GAS CONDITIONING
AND
PROCESSING

Volume 1: The Basic Principles

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VOLUME 1

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This book is one of a series in the area of petroleum production/processing. Each uses a CPS format designed to provide direct, concise explanations of the system involved for convenient use as a desk reference.

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ACKNOWLEDGMENT

This Seventh Edition represents a substantial rewrite and reorganization of the previous edition. This was done to reflect what I regard as necessary basic knowledge to function effectively in the subject area now and in the 1990's.

About thirteen revisions have occurred since this book had its birth as an adult training manual in 1960. Scores of folks have contributed to these changes by their formal or informal comments. Where possible, these have been cited. The source of some changes has been lost. I both apologize for this and thank these unknown persons for their contribution.

Special thanks are due Bob Hubbard, Larry Lilly, John Morgan, and Bob Maddox for their contributions in the preparation and review of the technical material. Sharon Antram is responsible for much of the new word processing and artwork. Joy Garrison (retired) is remembered for the many hours dedicated since 1968 to this and previous editions. Danny Stowe deserves special recognition for countless hours of layout and editing work. His efforts pulled this manuscript together as a book. Thank you all for your help.

Dr. John M. Campbell

Norman, Oklahoma
July 1992

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INTRODUCTION

"Knowledge of the basic fundamentals is the required foundation for successful professional practice."

This new Seventh Edition of "Gas Conditioning and Processing" has been edited to reflect continuing changes in our technology and the manner in which it is practiced. Most engineering calculations are performed on a computer and there often is little opportunity for the user to become very proficient in all areas for which he or she is responsible. The only practical answer is to understand the relatively few fundamentals which govern the multitude of applications for them.

As noted in the Campbell Rules of Engineering which follow, it is impossible to use a computer intelligently unless the user understands the principles governing the system in question. It is desirable to also have first-hand experience with that system, but that is not always feasible.

We continue to modify, delete, or add to the technical content, but the goal remains constant – providing reliable, credible basic information to enhance the ability of the individual professional. The emphasis is on relatively simple calculations and correlations that will give reliable answers for the applications involved. Companion computer programs are even available to make the book calculations less of a chore.

You may be somewhat surprised to find that some of these simple appearing approaches actually give more realistic results than some of the more complex simulation programs now available. This is not an accident or mere coincidence. Nothing goes into this book until it has been tested against the *actual* performance of the equipment or system involved, within the limit of conditions specified. For this reason, a primary use of these two volumes is to check the credibility of simulation calculations.

Another common use is to provide a "quickie" calculation for use in the early feasibility and planning of a project. Or, to troubleshoot and/or improve operating efficiency.

Simplicity has its virtues but it also imposes limitations. One should not extrapolate the correlations beyond their stated limits without concern for the increased uncertainty involved. Some of the factors affecting this are discussed throughout the book.

There are many figures and tables throughout this book that simply illustrate the relationship between the variables. They serve a kind of educational function to better understand how a piece of equipment works and to serve as a basis for the effective use of judgment – always a necessary ingredient in the engineering process.

Before addressing the details, I must admonish you to always remember that the professional decision process depends on factors that are not quantitative or calculable. These include economic, political, cultural, legal, and moral issues that may turn out to have more effect on the final decision than the technical aspects. We must always remember that engineering cannot be practiced in a vacuum.

The remainder of this chapter summarizes some of the terms and nomenclature that will be used throughout the book. One complication in this area is the inconsistency of numerical systems used throughout the world. It is interesting how otherwise perfectly rational persons will wax emotional about any deviation from the system they learned in school. So ... dual units are used in the hope that most users will be reasonably happy (even though not totally satisfied).

SOME CAMPBELL LAWS OF ENGINEERING

- "The utility of a computer calculation is limited by the user's real knowledge of the physical system involved and not by computer capability."
- "The time spent on a calculation tends to fully occupy the time available."
- "One cannot calculate reliably what one cannot measure reliably."
- "There is a tendency to concentrate effort on those calculations one knows how to make, using equations that happen to fit the data available and which yield explicit answers - not necessarily those that are important."
- "Many calculations have as their primary purpose the justification of pre-conceived opinions or positions (biases) rather than to obtain knowledge."
- "The true precision and/or accuracy of an answer can be no better than that of the worst input data; using a large number of significant figures cannot make a wrong answer right."
- "The use of statistics to handle uncertainty is a sound tool if used wisely but remember that statistics are like a bikini, what they reveal is interesting but that concealed may be vital."
- "Engineering calculations are somewhat like sex. They are important and give the practitioner varying amounts of pleasure, but there are other important factors to be considered in the total scheme of things."
- "Conventional wisdom can never be more than a partial truth because it is based on the past and assumes no change in the interim."
- "There is no such thing as a single equation or computer program that is equally valid for all mixtures in the full range of conditions under which they may exist."
- "The validity and usefulness of a computation is not necessarily proportional to its complexity."

The Fundamental Technology

All of the systems discussed in these two volumes are governed by a small group of fundamental laws and concepts. Understanding these, and their nomenclature, is critical. The most important are:

- The physical properties of the fluids and solids. (A fluid is anything that flows – liquid and/or gas.)
- The vapor-liquid-solid behavior of a substance as a function of pressure, temperature, and composition.
- The concept of equilibrium.
- The use of P-V-T equations of state to predict the behavior and properties of systems.
- The use of combination rules for the prediction of mixture behavior from that of individual molecular species present in the mixture.

- Thermodynamic concepts that govern the conservation of mass and energy, and the calculation of the energy level under given conditions.
- Rate equations which describe the rate of something happening in terms of stated driving and resisting forces, the area through which the transfer takes place, the configuration of the system, and a proportionality constant which covers all variables not specifically designated in the equation. This constant is most often called a "coefficient" with some adjective like heat or mass to designate the system it is used with.
- The molecular theory of substance behavior (indirectly).

In addition, it is required that the practitioner have reasonable literacy in the mathematical, chemical, and physical sciences.

Chapters 3-8 in Volume 1 and Chapter 12 in Volume 2 discuss the above in detail. Please note the discussions at the beginning of each of these chapters. The remainder of the chapters are applications of the aforesaid fundamentals.

Chapter 2 is a brief summary of system planning, oftentimes the most important aspect of the entire engineering effort. Each project is unique to some degree. When you decide to simply clone an existing system that is part of company standards or one with which you are familiar, without thorough review of the qualitative differences, you can be certain that at least a small mistake has been made.

Merely turning this phase over to an engineering company for a "feasibility study" also does not satisfy the need automatically. From their perspective, they will concentrate on simply modifying a process or system they know how to do or which they are licensed to use. Don't expect them to suggest you buy something that does not give them a competitive advantage. As an employee of the customer you have an ethical requirement that cannot be transferred (to another party – or a computer).

With these thoughts in mind, let us begin our journey through the book.

BASIC TERMINOLOGY

Throughout this book we will be using some basic chemistry and physics terminology. This is reviewed briefly.

Elements and Atoms

All matter in the universe is composed of *elements* which cannot be broken down or subdivided into smaller entities by ordinary means. Over 100 materials have been found (or created) which are classed as elements. These include carbon, hydrogen, sulfur, oxygen, nitrogen and helium – all materials occurring in petroleum systems.

The *atom* is the basic unit of each element that can combine with itself or the atoms of other elements to form a compound.

Chemical Compounds and Molecules

A true *compound* is a substance composed of more than one atom that satisfies both of the following conditions.

1. The atoms have combined chemically.
2. The compound formed possesses properties different from the atoms of which it is composed.

Chemical compounds are formed by the union of atoms in the simplest possible numerical proportions. The *molecule* is the unit of a compound. A molecule of water is H_2O, two atoms of hydrogen combined with one atom of oxygen.

A *diatomic molecule* is formed by the combination of two atoms of the same element. Nitrogen (N_2) and oxygen (O_2) are the most common examples.

Physical Compounds

A type of physical compound, called a *clathrate*, may be formed. A gas hydrate is one example of a clathrate. These compounds are relatively unstable.

Relative Atomic Mass (Weight)

In forming a compound, elements always combine according to fixed mass ratios. It is convenient to use relative atomic weight, the relative mass of the atoms of different elements, to express these ratios. The word "relative" means that the number used is a relative one. Currently, Carbon-12 is used as the standard, being assigned the relative atomic weight of 12. On this basis, the relative atomic weights of common oil and gas components are shown below (to the nearest whole number).

Atom	Symbol	Relative Atomic Weight
Hydrogen	H	1
Carbon	C	12
Nitrogen	N	14
Oxygen	O	16
Sulfur	S	32

A relative atomic weight of one element contains about the same number of atoms as a relative atomic weight of any other element. One gram of hydrogen and 12 grams of carbon each contain about 6×10^{23} atoms. Since relative weights represent a fixed number of atoms, they may be substituted for atoms in calculations.

Relative Molecular Mass (Weight)

The relative molecular weight of a molecule is the sum of the relative atomic weights of the atoms combining to produce the molecule. Water has a relative molecular weight of 18 ($H_2O = 2 + 16 = 18$). A diatomic molecule like oxygen (O_2) has a molecular weight of 32.

The Mol

The term "*mol*" is the historical abbreviation of the words "gram molecule." The current definition of the mol is: "The mol is the *amount of substance* of a system which contains as many *elementary entities* as there are atoms in 0.012 kilogram of Carbon-12." These elementary entities must be specified but include atoms, molecules, ions, electrons, etc.

The quantity 0.012 kg is 12 g, the relative atomic weight of carbon. Thus, the mol can be defined for engineering usage as that mass in grams equal numerically to the sum of the relative atomic weights of the atoms in the molecule of that substance.

By virtue of the definition, whenever the mol is used as a mass quantity without a prefix, a mass in grams is implied. If a relative molecular or atomic weight is expressed in pounds mass or kilogram, we will use the terms lb-mol and kmol, respectively, to denote this is not a standard mol. A *kilomol* (kmol) is simply 1000 mol.

$$1 \text{ kmol} = 1000 \text{ mol} = 2.205 \text{ lb-mol}$$

$$1 \text{ lb-mol} = 454 \text{ mol}$$

Throughout this book the *mol* will be used in many cases as a mass term in those processes where no chemical changes occur. It is particularly useful for gas calculations. At a given pressure and temperature equal volumes of different gas contain the same number of molecules. At 0°C [32°F] and 101.325 kPa [14.7 psia] a mol of any gas contains about 6×10^{23} molecules and occupies a volume of 22.4 liters.

The mol is thus a useful conversion factor from volume to mass, for the number of mols per unit volume is independent of gas composition.

Valence

Valence is a measure of the ability of atoms to form molecules by filling the electron shells of the atoms involved. The valence number is plus or minus, denoting the number of excess or shortage of electrons needed to fill its outer shell.

The question of atomic bonding is a complex subject involving many factors, as discussed in standard chemistry references. The concept is mentioned only to point out that the number of *bonds* or *linkages* used in the structural formulas that follow in the next section reflect the valence of the atoms in these compounds.

Mixture

A mixture is a combination of elements and compounds which may be separated by physical means. The properties of the mixture are a reflection of the properties of the constituents.

Natural gas and crude oil are mixtures. They are analyzed by separating the mixture into its component parts and identifying each by its properties.

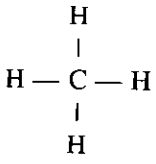
BASIC HYDROCARBON NOMENCLATURE

By definition, a hydrocarbon is any compound composed solely of carbon and hydrogen atoms. These atoms can combine in a number of ways to satisfy valence requirements. For convenience, these are separated into "families" or homologous series, each of which is given a name.

The carbon atoms can link together to form "chains" or "rings." Crude oil and natural gas mixtures consist primarily of "straight chain" hydrocarbon molecules, the bulk of which are paraffins.

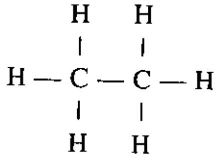
Paraffin Series Formula: C_nH_{2n+2}

Hydrocarbons in this series are *saturated* compounds – all four carbon bonds are connected either to another carbon atom or a hydrogen atom, with one such atom for each bond.



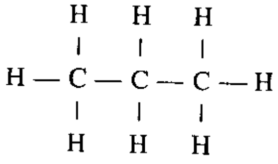
$$\begin{array}{c} H \\ | \\ H-C-H \\ | \\ H \end{array}$$

Methane



$$\begin{array}{c} H\ \ H \\ |\ \ \ | \\ H-C-C-H \\ |\ \ \ | \\ H\ \ H \end{array}$$

Ethane



$$\begin{array}{c} H\ \ H\ \ H \\ |\ \ \ |\ \ \ | \\ H-C-C-C-H \\ |\ \ \ |\ \ \ | \\ H\ \ H\ \ H \end{array}$$

Propane

Notice that all names end in *-ane*, the ending used for the paraffin series. In each case, the number of hydrogen atoms is two times the number of carbon atoms plus two more for the ends of the chain.

The paraffin hydrocarbons are the most stable of the lot because all valence bonds are fully satisfied as indicated by the single line linkage. Most reactions involve the replacement of hydrogen atoms with other atoms; the carbon linkage remains stable.

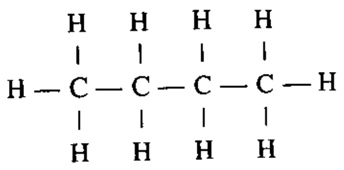
Each successive molecule in the paraffin series is created by adding a carbon and two hydrogens to the previous molecule. The incremental change in relative molecular weight is thus fourteen. Long chains containing scores of carbon atoms in series may be formed. However, the only ones normally identified by name contain ten or less carbons.

Name	Formula	Mol. Wt.	Name	Formula	Mol. Wt.
Methane	CH_4	16	Hexane	C_6H_{14}	86
Ethane	C_2H_6	30	Heptane	C_7H_{16}	100
Propane	C_3H_8	44	Octane	C_8H_{18}	114
Butane	C_4H_{10}	58	Nonane	C_9H_{20}	128
Pentane	C_5H_{12}	72	Decane	$C_{10}H_{22}$	142

In referring to a given paraffin hydrocarbon, the abbreviation C_3 for propane, C_4 for butane, etc. may be used. Statements like "propanes plus fraction (C_3+)" refer to a mixture composed of propane and larger atoms.

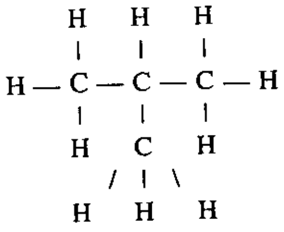
Paraffin Isomers. When the paraffin series molecule contains four or more carbon atoms there are different ways these can be connected without affecting the formula. Compounds which have the same chemical formula but a different atomic structure are called isomers. They possess different physical and chemical properties.

There are only two isomers of butane. In the structural diagram shown for i-butane we could draw the carbon atom below instead of above the carbon chain. But, this would be just a "mirror image" of the molecule as drawn. It is the same molecule with the same properties. The adjective "normal" is used to designate a molecule wherein all of the carbon atoms are in a straight line. An "isomer" has the same formula but a different arrangement of the carbon atoms. In an analysis, these are often abbreviated as "n" and "i" respectively.



$$\begin{array}{ccccccccccc} & & H & & H & & H & & H & & \\ & & | & & | & & | & & | & & \\ H & - & C & - & C & - & C & - & C & - & H \\ & & | & & | & & | & & | & & \\ & & H & & H & & H & & H & & \end{array}$$

Normal butane
(n-butane)



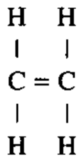
Isobutane
(i-butane)

Olefin or Ethylene Series (Alkenes) Formula: C_nH_{2n}

The olefin group of compounds is a simple straight chain series in which all the names end in *-ene*. Ethylene (ethene) C_2H_4 is the simplest molecule in the series.

Hydrocarbons in this series combine easily with other atoms like chlorine and bromine, without the replacement of a hydrogen atom. Since they are so reactive, they are called *unsaturated hydrocarbons*.

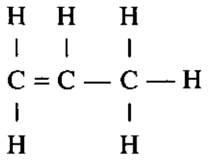
Unlike the paraffins, the maximum bonding capacity of the carbon atom is not fully satisfied by hydrogen or carbon atoms. Two adjacent carbon atoms form a "temporary" bond (in the absence of other available atoms) to meet bonding requirements fixed by valence. It is a necessary but unstable alliance. The structural formula for the olefins uses a double line to indicate the double carbon-carbon linkage, the most reactive point in the molecule.



$$\begin{array}{ccc} H & & H \\ | & & | \\ C & = & C \\ | & & | \\ H & & H \end{array}$$

, $H_2C=CH_2$

Ethylene (Ethene)

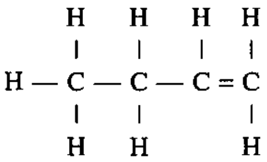


$$\begin{array}{ccccccc} H & & H & & H & & \\ | & & | & & | & & \\ C & = & C & — & C & — & H \\ | & & & & | & & \\ H & & & & H & & \end{array}$$

, $H_2C=CHCH_3$

Propylene (Propene)

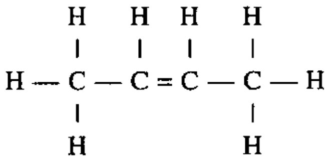
With four or more carbon atoms, isomers also may result from the position of the double bond as well as the arrangement of the carbon atoms.



$$\begin{array}{ccccccccc} & & H & & H & & H & & H \\ & & | & & | & & | & & | \\ H & — & C & — & C & — & C & = & C \\ & & | & & | & & & & | \\ & & H & & H & & & & H \end{array}$$

, $CH_3CH_2CH=CH_2$

1-Butene



$$\begin{array}{ccccccccccc} & & H & & H & & H & & H & & \\ & & | & & | & & | & & | & & \\ H & — & C & — & C & = & C & — & C & — & H \\ & & | & & & & & & | & & \\ & & H & & & & & & H & & \end{array}$$

, $CH_3CH=CHCH_3$

2-Butene

These molecules possess many different properties. They may furthermore react at the double bond or be split into two molecules at the double bond to form compounds with different characteristics.

The amount of olefins in natural gas usually is fairly small. Certain crude oils contain them in measurable amounts.

Acetylenic or Alkyne Series Formula: C_nH_{2n-2}

This series is of basic importance only in certain refining and petrochemical applications. Acetylene is the most important member of this series. It has the formula C_2H_2. The structural formula for acetylene is

$$H-C\equiv C-H$$

There is a triple bond between the carbon atoms. This satisfies the valence requirements but the carbon linkage is very weak.

Acetylene is even more reactive than the olefins. Carbon likes the sharing of three valence linkages even less than sharing two. Acetylene not only is unsaturated, it is almost unstable chemically. In the liquid state it is explosive if subjected to a sudden shock.

Diolefins Formula: C_nH_{2n-2}

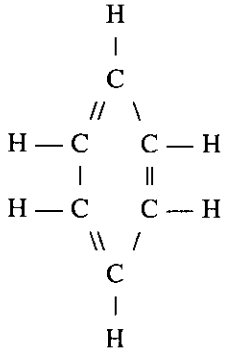
The *diolefins* have the same formula as acetylene. The members of this series contain two double linkages. They normally are named by replacing the *-ane* for paraffins by *-diene.*

Diolefins are primarily of concern in petrochemical plants. Butadiene is possibly the most interesting and useful since it is a primary ingredient in synthetic rubber compounds. It has the formula:

$$CH_2=CH-CH=CH_2$$

All of these unsaturated compounds are reactive. They may be hydrogenated. Liquid cooking oils (unsaturated) may be hydrogenated to form solid fats.

These compounds also polymerize – the process wherein a very large molecule is built up from the self-addition reaction of small identical molecules (monomers). Ethylene and propylene polymerize to form polyethylene and polypropylene, the basic ingredients in plastic materials. Acetylene polymerizes to form benzene, a cyclic hydrocarbon.



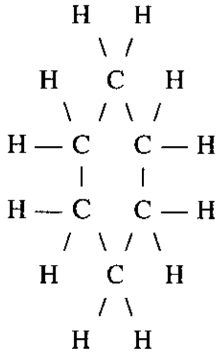
Benzene

Aromatic (Benzene) Series Formula: C_nH_{2n-6}

Aromatic is the word used to describe an unsaturated hydrocarbon molecule where the carbon atoms form a ring, a cyclic compound. *Benzene*, the parent compound of this series, has the structural formula of C_6H_6.

Since the aromatics are *unsaturated*, they react readily. They may be oxidized to form organic acids. They also promote foaming and other operational problems in the production and handling of crude oil and natural gas.

Most petroleum contains only a trace of aromatics. Some contain significant amounts. Any analysis of crude oil and natural gas should include aromatics. Even small amounts can influence physical behavior and affect design.



Cyclohexane

Naphthene Series Formula: C_nH_{2n}

The naphthene series has a ring structure but is saturated. Naphthenes may be found in most crude oils but are seldom shown in routine analyses. Being saturated molecules, they are not very reactive. Cyclohexane is a common member of this series. Its structural formula is C_6H_{12}.

Cyclohexane is similar to benzene except that it is saturated. On chromatographic analysis it occurs between n-hexane and n-heptane. Cyclopentane (C_5H_{10}) also occurs. On chromatographic analysis it occurs between n-pentane and n-hexane.

PARAFFIN HYDROCARBON COMPOUNDS

In the production, gathering, conditioning, and processing of natural gas and its associated liquids, the primary concern is the behavior of the paraffin series hydrocarbons with 10 or less carbon atoms (C_1–C_{10}). This concern includes nitrogen and water and contaminants in the gas, such as sulfur compounds.

Paraffin hydrocarbons are less reactive with other materials than many hydrocarbons, but it must be remembered that they have been in contact with the chemicals present in the reservoir rock for many millions of years. They are also conditioned by use of alcohols, glycols, and amines in which they are soluble and with which they react to some degree.

Radicals

A radical represents a group of atoms that act as a single unit in the formation of many common compounds.

Alkyl Radical. At least the simpler paraffins often react by replacing one hydrogen with some other radical or element. This alkyl radical has the formula:

$$C_nH_{2n+1} : \underset{\text{methyl}}{(CH_3)},\ \underset{\text{ethyl}}{(C_2H_5)},\ \underset{\text{propyl}}{(C_3H_7)}$$

The parenthesis indicates the radical group. The alkyl radical normally has a valence of +1.

In many cases the alkyl radical is indicated by the symbol "R." The formula for methanol is CH_3OH; for ethanol it is C_2H_5OH. Both may be written as ROH. When "R" is used, one cannot identify the specific alkyl radical. It is used only to show general reaction characteristics.

Hydroxyl Radical, (OH). This combination occurs in many common compounds. It combines with hydrogen to form water – H(OH) or H_2O; with metallic salts like sodium, calcium and magnesium to form hydroxides (bases, caustics); and with alkyl radicals to form alcohols, such as methanol, ethanol, etc.

(SO_4), (CO_3). If radicals like these combine with hydrogen, an acid is formed. When combined with metallic salts like sodium, calcium, and magnesium, a salt is formed (which occurs commonly in water systems). The scale formed in water systems is caused by precipitation of salts like these. The common names for some common radicals of this type are:

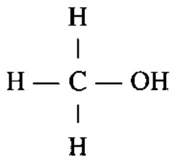
SO_4 – sulfate CO_3 – carbonate

SO_3 – sulfite HCO_3 – bicarbonate

Each of the radicals has a valence found from the valence of its elements. The hydroxyl radical (OH) has a valence of minus one and is sometimes written as $(OH)^{-1}$. It therefore combines in proportions fixed by this valence: H(OH), NaOH, $Mg(OH)_2$ – so that the sum of plus and minus valences equals zero.

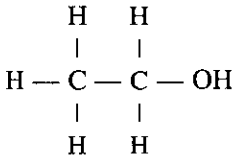
Alcohols

The common alcohols are formed from the addition of a single hydroxyl radical to an alkyl radical. The name of the alcohol ends in "ol," or the name of the alkyl radical is followed by the word "alcohol."



$$\begin{array}{ccccc} & & H & & \\ & & | & & \\ H & — & C & — & OH \\ & & | & & \\ & & H & & \end{array} \quad , \quad CH_3OH$$

Methanol or methyl alcohol



$$\begin{array}{ccccccc} & & H & & H & & \\ & & | & & | & & \\ H & — & C & — & C & — & OH \\ & & | & & | & & \\ & & H & & H & & \end{array} \quad , \quad C_2H_5OH$$

Ethanol or ethyl alcohol

Both C_2H_5OH and CH_3OH could be written as ROH in denoting the general reaction of an alcohol.

Mercaptans

Compounds with the general formula RSH are known as mercaptans. They may be regarded as sulfur alcohols since the formula is the same if you replace the oxygen atom in the (OH) radical by a sulfur atom.

Formulas for typical mercaptans are:

CH_3SH – methyl mercaptan C_2H_5SH – ethyl mercaptan

Other Carbon-Sulfur Compounds

There are several other carbon-sulfur compounds present in sour petroleum fluids. Some are:

Carbonyl sulfide – COS   Carbon disulfide – CS_2

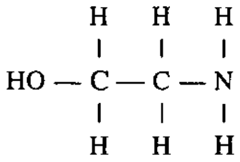
Thiophene – an unsaturated compound having the formula: HC = CH – S – HC = CH

Sulfur is a very reactive element that combines chemically with many other elements and compounds. Its compounds react with carbon steel to form sulfides and oxides of iron. Many compounds polymerize and form the "sludge" so common in sour petroleum systems. This sludge is often very corrosive and should be removed by filtration.

Organic Nitrogen Compounds - Amines

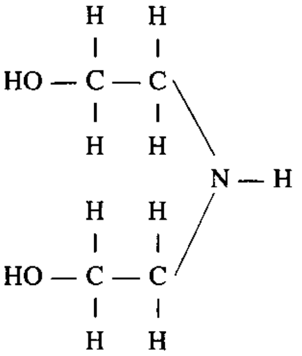
There are a number of common organic compounds formed by the reaction of organic materials with ammonia (NH_3). In this basic reaction one or more hydrogen atoms are replaced by an organic radical. The word "amine" is used commonly to denote this type of compound. There are a large number of amines used in the chemical industry.

The alkanolamines are commonly used for treating sour gases and liquids, particularly monoethanolamine and diethanolamine. As the names indicate, the alkanolamines may be considered a combination of an alcohol and ammonia.



, $HOCH_2CH_2NH_2$

Monoethanolamine (MEA)



, $(HOCH_2CH_2)_2NH$

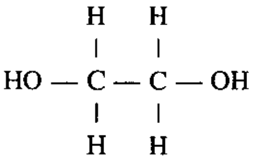
Diethanolamine (DEA)

Notice that the only difference in the above compounds is how many hydrogen atoms of ammonia are replaced by the radical (C_2H_4OH), ethanol minus one hydrogen atom.

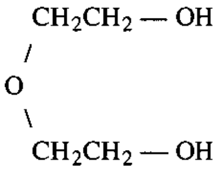
Glycols

The glycols are a family of chemicals, sometimes called diols. They may be regarded as complex alcohols since they contain alkyl and hydroxyl radicals. The glycols used for dehydration are based on the ethyl radical.

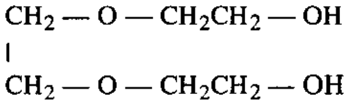
As with most compounds containing hydroxyl groups, the glycols react readily with other compounds and elements. In DEG and TEG the oxygen atom also is very reactive.



Ethylene Glycol (EG)



Diethylene Glycol (DEG)



Triethylene Glycol (TEG)

ACIDS, BASES AND SALTS

We are always talking about acids and bases in the handling of petroleum-water mixtures, processing, and other such functions of the petroleum industry.

Except for water and hydroxides (compounds containing the OH radical), all inorganic compounds of hydrogen are acids. They consist of hydrogen combined with an acid radical (anion).

Acid Radical (anion)	Symbol	Acid	Formula
Chloride	Cl^{-1}	Hydrochloric	HCl
Carbonate	CO_3^{-2}	Carbonic	H_2CO_3
Sulfate	SO_4^{-2}	Sulfuric	H_2SO_4
Nitrate	NO_3^{-1}	Nitric	HNO_3
Phosphate	PO_4^{-3}	Phosphoric	H_3PO_4

Notice that hydrogen (valence of +1) combines with the acid radical in a proportion such that the net valence of the compound formed is zero. This is the rule to be followed in all compound formation. Since the valence of the sulfate radical is -2, it takes two hydrogens.

The combination of a metal cation such as sodium with the hydroxyl anion (OH) produces a base. Sodium hydroxide (NaOH) is commonly called caustic.

pH. The acidity or alkalinity of a material is measured on a scale similar to that of a thermometer. This pH scale is the logarithm of the reciprocal of the hydrogen ion concentration. It runs between 0 and 14. A pH of 7 is neutral. Acids have a pH less than 7; bases (alkaline solutions) have a pH greater than 7.

Since pH is a logarithmic function, a solution possessing a pH of 5.0 is 100 times more acidic than one with a pH of 7.0.

ANALYSIS OF MIXTURES

A routine analysis of a hydrocarbon mixture is shown in Table 1.1. Notice that only paraffin hydrocarbons are shown. This is not entirely correct, although the paraffins may be the predominant series present. Notice also that all molecules heptane and larger are lumped together as a heptanes plus fraction.

The hydrocarbon portion of an analysis like this usually is obtained from a chromatograph. The printed output from this technique is a series of "peaks" rising from a base line. The area under the peak for any component is proportional to the amount present. The instrument is calibrated using standard samples of known analysis so that peak area can be converted to the amount present.

Figure 1.1 is an example of a chromatogram (chart) from a chromatographic analysis. When the ordinary peak height is so low it is difficult to measure, it is attenuated (multiplied). The attenuation factor is used to convert the peak height shown to analysis.

The chromatograph used to obtain Figure 1.1 was capable of showing hydrocarbons other than paraffins. Notice that all of these occur in the hexane-heptane range of molecules. Many chromatographs, particularly those used for routine gas analyses, are unable to detect all of these components. The nonparaffins combine with a single peak which is reported (erroneously) as being a paraffin material.

Many system operating problems are the direct or indirect result of inadequate analysis. This may result from failure to:

1. Analyze for CO2 and sulfur compounds
2. Identify the presence of aromatics and other nonparaffin hydrocarbons
3. Adequately characterize the heaviest fraction

TABLE 1.1
Fluid Analysis

Component	Mol %
Nitrogen	0.34
Carbon Dioxide	0.91
Sulfur (as H_2S)	0.50
Methane (C_1)	21.36
Ethane (C_2)	36.78
Propane (C_3)	10.21
i-Butane (iC_4)	6.38
n-Butane (nC_4)	9.84
i-Pentane (iC_5)	2.63
n-Pentane (nC_5)	4.01
Hexanes (C_6)	3.90
Heptanes Plus (C_7+)	3.14
Total	100.00
Mol wt. C_7+ = 118	
Rel. Density C_7+ = 0.71	

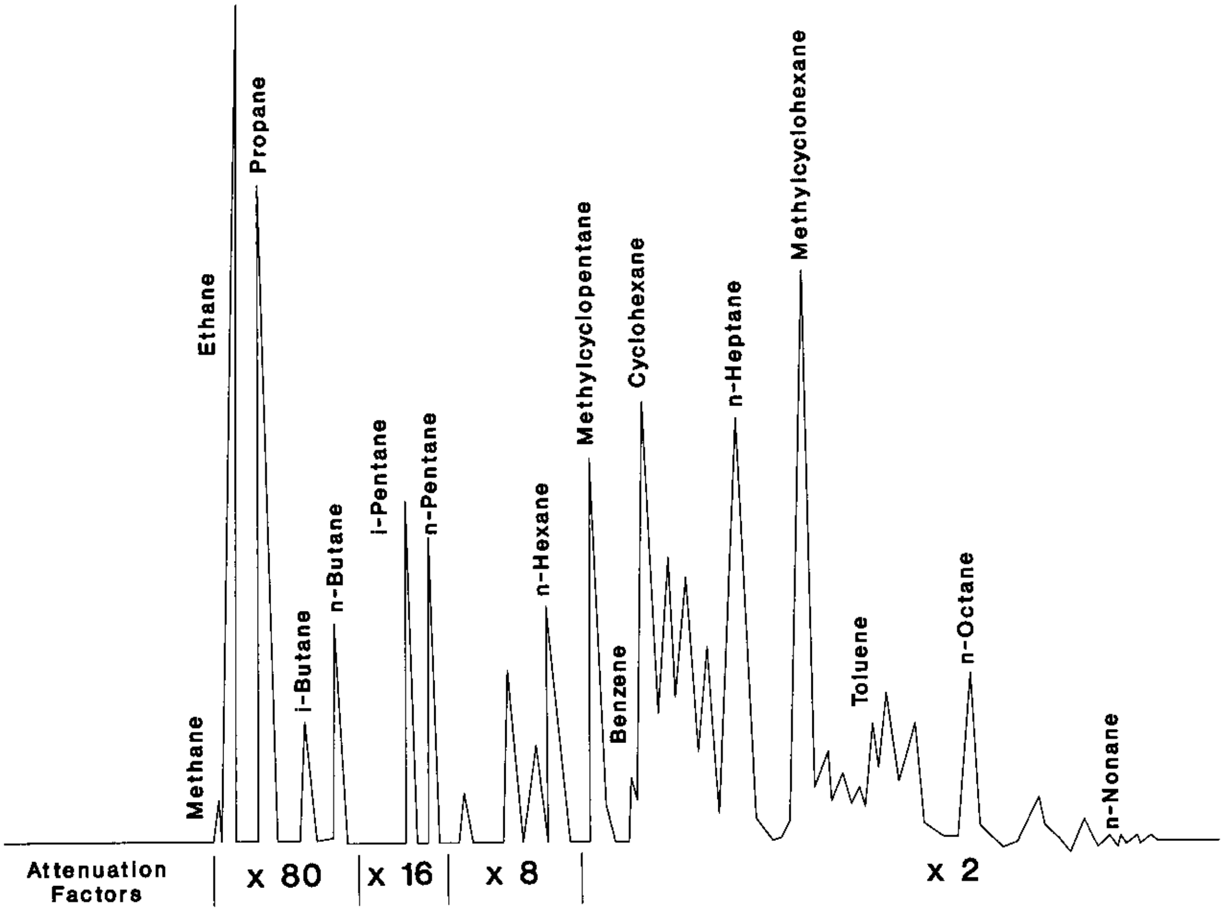


Figure 1.1 **Chromatogram of Condensed Liquid**

All of these will be discussed in more detail in subsequent chapters. There are, however, some general guidelines.

1. Always analyze for CO_2 and H_2S. Sulfur contents as low as 3-10 parts per million may prove troublesome. If the sulfur content (reported as H_2S) is higher than this, a special analysis for carbonyl sulfide (COS), carbon disulfide (CS_2) and mercaptans is intelligent.
2. Some crude oil may contain up to 10-12% aromatics. Failure to be aware of this affects mechanical design problems and reduces validity of equilibrium calculations.
3. If the gas is from a separator, characterizing the heaviest fraction through C_{7+} may be adequate; if it is wellbore gas it may not be adequate. Crude oil may need to be characterized through C_{20+} to achieve reliable equilibrium predictions.

The prompt and proper analysis of representative samples is a critical factor. Any calculation is an exercise in futility unless the analyses used are reliable. Inadequate sampling and analysis is a major cause of problem systems.

A sampling program should be planned carefully. Appendix A, at the end of this volume, summarizes good sampling procedures.

BASIC UNITS OF MEASUREMENT

Dual English/Metric units are used throughout. This is a necessary compromise at this point in time because the U.S. petroleum industry still uses English units and the progress toward change is relatively slow. Since it is not economically feasible to publish separate editions, dual units are used. There is some element of frustration involved, but we have found that dual units help make the user more "bilingual," a useful trait.

Metric units of the *System Internationale* (SI) are used. The basic definitions and practices under this system have been standardized by interlocking international bodies, but the application of these is not completely uniform. Herein we will use the standards of the Society of Petroleum Engineers (SPE), the primary professional society, international in scope, serving the petroleum industry.*(1.1)*

Appendixes 1A and 1B at the end of this chapter define the unit definitions developed for the SI metric system. Figure 1.2 shows the basic relationships between these units.

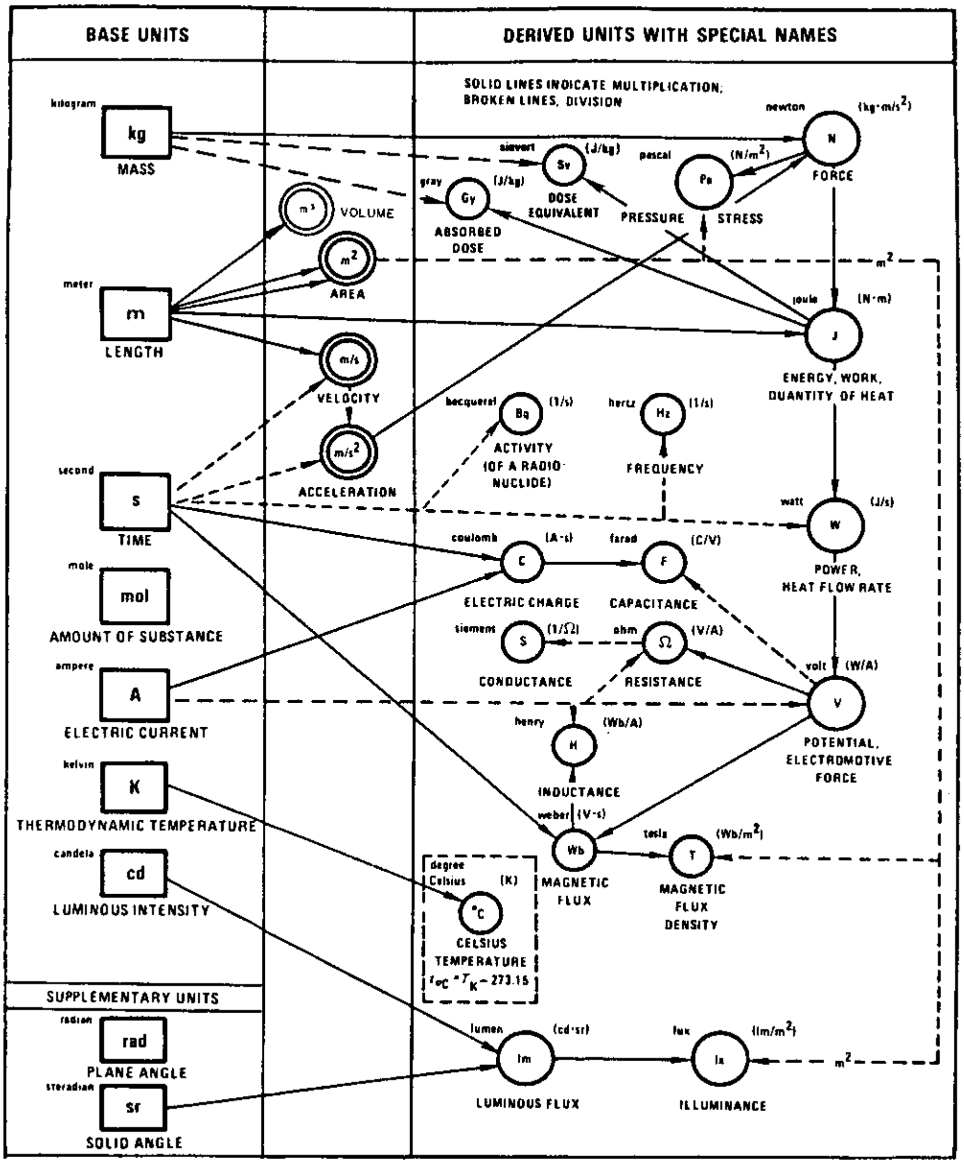


Figure 1.2 Relationship between SI Metric Units

The SI system possesses many advantages. It is not ambiguous; only one unit is defined for each quantity. Each unit also possesses a unique, carefully defined symbol. A decimal (powers of ten) relationship exists between multiples of units. The system of units also has been chosen in such a way that equations between numerical values and quantities both possess the same form. In a coherent system like this no conversion numbers are needed in the equations so long as one expresses all quantities in basic or derived units. The conversion number always is unity.

However, it really is not very feasible to express all quantities in basic or derived units, particularly when dealing with pressure and energy. Very, very large or small numbers result, which are inconvenient to handle arithmetically or difficult to compare. Thus, a series of prefixes has been approved, as shown in Table 1.2.

TABLE 1.2
Summary of SI Metric Prefixes

Multiplication Factor	SI Prefix	SI Prefix Symbol, Use Roman Type	Pronunciation (US)	Meaning (US)	Meaning in Other Countries
1 000 000 000 000 000 000 = 10^{18}	exa	E	ex' a(*a* as in *a*bout)	one quintillion times	trillion
1 000 000 000 000 000 = 10^{15}	peta	P	as in *pet*al	one quadrillion times	thousand billion
1 000 000 000 000 = 10^{12}	tera	T	as in *terra*ce	one trillion times	billion
1 000 000 000 = 10^{9}	giga	G	jig'a (*a* as in *a*bout)	one billion times	milliard
1 000 000 = 10^{6}	mega	M	as in *mega*phone	one million times	
1 000 = 10^{3}	kilo	k	as in *kilo*watt	one thousand times	
100 = 10^{2}	hecto	h	heck'toe	one hundred times	
10 = 10	deka	da	deck'a (*a* as in *a*bout)	ten times	
0.1 = 10^{-1}	deci	d	as in *deci*mal	one tenth of	
0.01 = 10^{-2}	centi	c	as in *senti*ment	one hundredth of	
0.001 = 10^{-3}	milli	m	as in *mili*tary	one thousandth of	
0.000 001 = 10^{-6}	micro	µ	as in *micro* phone	one millionth of	
0.000 000 001 = 10^{-9}	nano	n	nan'oh (*an* as in *ant*)	one billionth of +	milliardth
0.000 000 000 001 = 10^{-12}	pico	p	peek' oh	one trilionth of +	billionth
0.000 000 000 000 001 = 10^{-15}	femto	f	fem' toe (*fem* as in *fem*inine)	one quadrillionth of	thousand billionth
0.000 000 000 000 000 001 = 10^{-18}	atto	a	as in an*atom*y	one quintillionth of	trillionth

Experience has shown that the use of not over 3 to 4 significant figures is both desirable and realistic. Prefixes are thus chosen to stay within this guideline. In a few selected cases, alternate units are allowed to minimize the need for prefixes.

Consistent with this, conversion factors herein are rounded off. This policy is compatible with the accuracy of data being addressed by the calculation.

Of course, there are cases with large or small numbers where the zeros are significant. To limit their use, one of the conventions shown below will be used.

Presenting Numbers

The traditional use of commas and periods has been changed. A comma or period is used only to denote a decimal point. For a number with more than four digits a space is now left where a comma or period was used to denote groups of three numbers. Either of these power forms is acceptable.

NEW	OLD
1 200 000.00	1,200,000.00 or 1.200.000,00
1200.00 or 1 200.00	1,200.00 or 1200.00 or 1.200,00
0.104	0.104 or 0,104
0.001 04	0.00104 or 0,00104

Numbers containing many zeros as significant figures are written conveniently as follows:

$$1\ 200\ 000 = 1.2 \times 10^{6} = 1.2(\mathrm{E}{+}06)$$

$$0.001\ 04 = 1.04 \times 10^{-3} = 1.04(\mathrm{E}{-}03)$$

Mass, Force and Weight

The mass of a body varies with its velocity but the velocity effect is trivial under the conditions discussed in this book. Thus, mass is considered constant. Mass as a quantity is a measure of the amount of matter in a body by virtue of which it possesses inertia. The basic units of mass used most commonly (pound and kilogram) are specified as the name for the mass of specific bodies of platinum iridium preserved as standards.

Force is the result when a mass is accelerated. This acceleration may be due to gravitational effects or imposed artificially. The basic relationship between mass and force is:

$$F = \frac{ma}{g_c} \tag{1.1}$$

Where: m = mass

a = acceleration

g_c = proportionality constant for force

One of the major differences between SI metric and the earlier metric systems is the manner in which the force-mass relationship is handled. Table 1.3 shows five systems that have been used. Column (5) shows the SI metric system units.

Note that the newton (N) is the force required to accelerate one kilogram at the rate of one meter per second squared. This is a coherent equation since the proportionality constant is unity.

In using Equation 1.1, note that the name for the force unit is governed by the mass unit, the acceleration rate specified and the numerical value of the proportionality constant. In most applications of this equation gravitational acceleration is used.

Weight is a term to be avoided even though it is often used to denote mass. Weight is a measurement of the attraction of gravity on a given mass. It thus varies with the magnitude of gravity. Weight is really a force.

The differentiation between weight and force has been muddled by the definition of a pound force (lbf) as a pound of mass (lbm) subject to the earth's gravitational acceleration of 32.2 ft/s^2 at 45° latitude and sea level. By setting the proportionality constant equal to this acceleration, a pound force (weight) was equal numerically to a pound mass. In the SI metric system no such ambiguity exists. The basic unit of force is the newton (N) and the basic unit of mass is the kilogram (kg).

TABLE 1.3

Force and Mass Units

	(1)	(2)	(3)	(4)	(5)
Mass	lbm	lbm	kg	g	kg
Length	ft	ft	m	cm	m
Time	s	s	s	s	s
Force	lbf	poundal*	kgf	dyne	N
a	32.17 ft/s^2	1.0 ft/s^2	9.81 m/s^2	981 cm/s^2	1.0 m/s^2
g_c	$32.17\,\frac{\text{lbm–ft}}{\text{lbf–f}^2}$	$1.0\,\frac{\text{lbm–ft}}{\text{pdl–s}^2}$	$1.0\,\frac{\text{kg·m}}{\text{kgf·s}^2}$	$1.0\,\frac{\text{g·cm}}{\text{dyne·s}^2}$	$1.0\,\frac{\text{kg·m}}{\text{N·s}^2}$

*Poundal is the name given to the unit ft-lb/s^2.

This standard acceleration = 32.2 ft/s^2 = 9.81 m/s^2. It is the one used normally to convert force to pressure.

Length, Area and Volume

The standard unit of *length* is the meter (m). Common length units employed are the millimeter (mm), centimeter (cm), decimeter (dm) and kilometer (km).

1 m = 1000 mm = 100 cm = 10 dm = 0.001 km
1 m = 3.28 ft, 1 ft = 0.305 m
1 statute mile = 1.61 km, 1 km = 0.621 statute mile

Area is expressed in the same length units squared.

1.0 m^2 = 1.0(E+06) mm^2 = 100 dm^2 = 10.76 ft^2
1.0 hectare(ha) = 1.0(E+04) m^2 = 2.47 acres
1.0 sq mile = 2.59 km^2 = 259 hectares

The basic *volume* term is cubic meters. The liter (L) is an allowable name for the cubic decimeter. The API barrel (bbl), which is 42 U.S. gallons, is not a part of the SI metric system.

1 m^3 = 1000 dm^3 = 1000 L = 35.31 ft^3
1 L = 0.001 m^3 = 1000 cm^3 = 0.0351 ft^3 = 61 in.^{3}
1 ft^3 = 0.0283 m^3 = 28.3 L = 7.48 U.S. gal. = 6.23 U.K. gal.
1 bbl = 0.159 m^3 = 159 L = 5.61 ft^3 = 42 U.S. gal. = 35 U.K. gal.

Density

The usual units are kg/m^3 or lbm/ft^3. Density varies with pressure and temperature. Standard or reference density usually is reported at standard atmospheric pressure and some temperature like 0°C, 15°C, 25°C, 60°F and 77°F. For most engineering calculations involving gross amounts of liquid, the standard density is good enough. For custody transfer and gases, an actual density is necessary at the P and T of the calculation.

1 kg/m^3 = 0.001 g/cm^3 = 0.0624 lb/ft^3
1 lb/ft^3 = 16.02 kg/m^3 = 0.016 g/cm^3
1 g/cm^3 = 1000 kg/m^3 = 1.0 kg/L = 62.4 lb/ft^3

The term (g/g_c) sometimes is called a "force equivalent" and is represented by the symbol F_e. The value will vary with the actual gravitational acceleration at the site, but the use of standard values normally is satisfactory.

Temperature

The SI units are degrees Celsius (°C) and Kelvin (K). The word "centigrade" is obsolete.

$$°C = 0.556\,(°F - 32), \quad K = °C + 273 \qquad °F = (1.8)(°C) + 32, \quad °R = °F + 460$$

Degrees Fahrenheit and Rankine are the corresponding values in the traditional English system of units. Absolute temperature is used in most engineering calculations.

Work and Power

The units of work are force times a distance. Power is the time rate of doing work. The standard work units are the joule (J) = 1.0 N·m and the ft-lbf. Once again the joule is a very small unit so the kilojoule (kJ) or megajoule (MJ) normally is used.

1 kJ = 0.001 MJ = 737 ft-lbf

Normally we are concerned with power. The watt (W) and horsepower (hp) are the most common terms.

1.0 W = 1.0 J/s, 1.0 hp = 550 ft-lbf/s
1.0 kW = 3600 kJ/h = 1.34 hp
1.0 hp = 0.746 kW = 2686 kJ/h

Heat Energy

As noted in later chapters, heat energy must have the equivalent units as work and power in order to make calculations involving the conversion between them. In SI metric the joule is used also for heat energy; the British Thermal Unit (Btu) is the traditional counterpart. There are several definitions of the Btu, but the International Table value will be used for all conversions. For our purposes the differences are trivial.

1.0 Btu = 1.055 kJ = 778 ft-lbf
1.0 kW = 3600 kJ/h = 3412 Btu/h
1.0 hp = 2545 Btu/h = 2686 kJ/h

Energy per unit mass will be expressed in kJ/kg and Btu/lbm, in most instances.

1.0 Btu/lbm = 2.326 kJ/kg = 6.46(E–04)(kW·h)/kg

Viscosity

The coefficient of viscosity (normally shortened to dynamic viscosity) is a shear force per unit area. The *poise* was the original unit; 1 poise = 1 (dyne-sec)/cm^2. For petroleum systems the centipoise (cp) is the most convenient unit. It may be expressed in force units or equivalent mass units. In SI metric the Pa·s is the preferred unit.

$$1.0 \text{ cp} = 0.001 \text{ Pa·s} = 0.001 \text{ kg/(m·s)} = 6.72 \times 10^{-4} \text{ lbm/(ft-s)}$$

All calculations in this book use dynamic viscosity. However, dynamic viscosity divided by density is known as *kinematic viscosity* with the primary unit being the centistoke (cSt).

$$1.0 \text{ cSt} = 1.0 \text{ mm}^2/\text{s} = 1.08(\text{E–05})\text{ft}^2/\text{s}$$

Relative density is the density of the system divided by that of a reference substance at specified conditions. Relative density is the new SI metric term replacing specific gravity. The reference for liquid is water; for gas it is air.

$$\gamma_L = \frac{(\ \rho\ \text{liquid})}{\underset{\text{water}}{(\ \rho\ \cancel{\text{liquid}})}} \qquad (1.2)$$

$$\gamma_g = \frac{(\rho\ \text{gas})}{(\rho\ \text{air})} = \frac{(\text{M gas})}{(\text{M air})}$$

Where: M = relative molecular weight

In solving Equation 1.2, the density of water usually can be taken as 1000 kg/m^3, 1.0 g/cm^3 or 1.0 kg/L; M of air is taken as 28.96 (or 29).

The relative density of an oil may be expressed in °API. The equation is

$$\gamma_L = \frac{141.5}{(131.5 + {}^\circ\text{API})} \qquad (1.3)$$

$$\text{API} = \frac{141.5}{\gamma_L} - 131.5$$

Pressure

Pressure is the force exerted per unit area. The standard pressure term is the pascal (Pa) = 1 N/m^2. This is a small pressure unit, 1 lbf/in.^{2} (psi) = 6895 Pa. A pressure of 500 psi would be 3 447 500 Pa, an awkward number to handle and deal with relatively. So, the kilopascal (kPa) and megapascal (MPa) are used most commonly. The use of the term kgf/cm^2 as a measure of pressure virtually has disappeared. In some areas the term bar is used as a special name for 100 kPa.

Absolute pressure is used in most engineering calculations. In English units we add an "a" to indicate absolute pressure (psia, psfa); "g" is added for gauge pressure (psig, psfg). In SI metric no suffix will be used for absolute pressure. A statement ***P = 138 kPa implies an absolute pressure***. If a gauge pressure is used it will be written P = 37 kPa(g).

100 kPa = 1 bar = 0.987 atm = 1.02 kgf/cm^2 = 14.50 psi
14.696 psia = 1 atm (std) = 101.325 kPa = 1.01 bar = 1.03 kgf/cm^2

For the purpose of engineering calculations, the difference between the use of 100 kPa and 101.325 kPa as an atmospheric pressure standard is trivial. All data being used have an accuracy less than any error involved.

Fluid Head. A vertical column of gas or liquid will exert a pressure on the bottom. This *head*, expressed in feet or meters, may be converted to a pressure term by the equation

$$P = A\,(\rho)\,(H)\,(g / g_c) \qquad (1.4)$$

Where:	Metric	English
ρ = fluid density	kg/m^3	lb/ft^3
H = fluid column	m	ft
g = grav. acceleration (std)	9.81 m/s^2	32.2 ft/sec^2
g_c = prop. constant	1.0 kg·m/N·s^2	32.2 lbm–ft/lbf–sec^2
P = pressure	kPa	psi
A = unit conversion factor	0.001	0.0069

VOLUMETRIC-MASS CONVERSIONS

The standard cubic foot and the standard cubic meter are a "bastard legacy" of historical contracts for the metering and custody transfer of natural gas. In engineering calculations, it is necessary to convert these volume terms to an appropriate mass term. This is routine but a nuisance.

This standard volume was designated at some prescribed base pressure and temperature. In the U.S. the historical base temperature has been 60°F; the historical base pressure 14.4-15.025 psia, depending on the organization. In recent years 14.696 psia, the standard atmospheric pressure, has been the most common pressure used.

There has been a lot of trivial controversy in former English unit areas about what comparable S.I. units to use. It is my position that the justification for metricating is its consistent "multiples of ten" character and corresponding coherence; one need not memorize a lot of odd numbers. One segment wishes to use 15°C since it is close to 60°F and 101.325 kPa, the equivalent of 14.696 psia. and. 23.6 m^3/kmol

Of course it makes little difference in engineering calculation accuracy which base one uses for conversion. However, I reject the argument that proposes odd numbers as being spurious and use standards that follow, in accordance with SPE practice.

Nm^3 0°C
101.355 kpa
224 m^3/kmol

English: $P_{sc} = 14.7$ psia, $T_{sc} = 60°F = 520°R$

Metric: $P_{sc} = 100$ kPa (1 bar), $T_{sc} = 15°C = 288$ K

The following basic conversion factors apply at these conditions.

English:	379.3 ft^3/lb-mol, 2636 lb-mol/MMscf	Density air = 0.0764 lb/ft^3
Metric:	23.96 m^3/kmol, 41 740 kmol/10^6 std m^3	Density air = 1.21 kg/m^3

In English units, M is the Roman numeral for one thousand; thus MM is one million. MMscf, MMSCF and MMCF are alternate ways of writing "one million standard cubic feet." In the metric system, "M" stands for "mega," the prefix for 10^6 or (E+06). To avoid confusion, millions of standard cubic meters will be written out or abbreviated as 10^6 std m^3.

The subscript "sc" always means standard conditions. So, $q_{sc} = 1.6 \times 10^6\ m^3/d$ means a flow rate of 1.6 million standard cubic meters per day. The abbreviation "std" is not required if the "sc" subscript is used as shown. One may not use any subscript on a metric unit; m^3_{sc} is forbidden by SI standards.

In the most common conversion, a volumetric flow rate per day is converted to a mass rate per hour. The use of lb-mol or kmol often is a convenient mass unit if no chemical reaction is occurring.

Gas:

$$\text{lb-mol/hr} = \left|\frac{\cancel{\text{MMscf}}}{\cancel{\text{day}}}\middle|\frac{\cancel{\text{day}}}{24\text{ hr}}\middle|\frac{2636\text{ lb-mol}}{\cancel{\text{MMscf}}}\right| = 110\ (\text{MMscf/d})$$

$$\text{lb/hr} = (\text{lb-mol/hr})(\text{M gas}) = 110\ (\text{MMscf/d})(\text{MW gas})$$

$$= \left|\frac{\cancel{\text{MMscf}}}{\cancel{\text{day}}}\middle|\frac{\cancel{\text{day}}}{24\text{ hr}}\middle|\frac{76\,400\text{ lb air}}{\cancel{\text{MMscf}}}\middle|\frac{\gamma_g}{}\right| = 3180\ (\text{MMscf/d})(\gamma_g)$$

$$\text{kmol/h} = \left|\frac{\cancel{10^6\text{ std m}^3}}{\cancel{\text{d}}}\middle|\frac{\cancel{\text{d}}}{24\text{ h}}\middle|\frac{41\,740\text{ kmol}}{\cancel{10^6\text{ std m}^3}}\right| = 1739\ (10^6\text{ std m}^3/\text{d})$$

$$\text{kg/h} = (\text{kmol/h})(\text{M gas}) = 1739\ (10^6\text{ std m}^3/\text{d})(\text{MW gas})$$

$$= \left|\frac{\cancel{10^6\text{ std m}^3}}{\cancel{\text{d}}}\middle|\frac{\cancel{\text{d}}}{24\text{ h}}\middle|\frac{1.21(\text{E+06})\text{ kg air}}{\cancel{10^6\text{ std m}^3}}\middle|\frac{\gamma_g}{}\right| = 50\,400\ (10^6\text{ std m}^3/\text{d})(\gamma_g)$$

Liquid:

$$\text{lb–mol/hr} = \left| \frac{\cancel{\text{bbl}}}{\cancel{\text{day}}} \middle| \frac{\cancel{\text{day}}}{\text{24 hr}} \middle| \frac{\text{350 lb } H_2O}{\cancel{\text{bbl}}} \middle| \frac{\gamma_L}{\text{M}} \right| = 14.6\ (\text{bbl/d})(\gamma_L/\text{MW})$$

$$\text{lb/hr} = (\text{lb–mol/day})(\text{MW}) = 14.6\ (\text{bbl/d})(\gamma_L)$$

$$\text{kmol/h} = \left| \frac{\cancel{m^3}}{\cancel{d}} \middle| \frac{\cancel{d}}{\text{24 h}} \middle| \frac{\text{1000 kg } H_2O}{\cancel{m^3}} \middle| \frac{\gamma_L}{\text{M}} \right| = 41.7\ (m^3/d)\ (\gamma_L\ /\text{MW})$$

$$\text{kg/h} = (\text{kmol/h})(\text{M gas}) = 1739\ (10^6 \text{ std } m^3/d)(\text{MW gas})$$

These and other conversion factors used are summarized in Appendix 1C at the end of this chapter. In making these conversions there are some small "round-off" deviations that have no significance in the calculations wherein they are used. Appendix B at the end of this volume is a more complete listing of conversion factors.

In using conversion factor summaries, which are presented differently, remember the following relationships.

Unit A times Conversion Factor B = Unit C

(ft)(12) = inches, (kg)(1000) = g

or One Unit of A = Conversion Factor B times Unit C

1.0 ft = 12 inches, 1.0 kg = 1000 g

It is easy to get the conversion backward, particularly when using unfamiliar units.

REFERENCES

1.1 "The SI Metric System of Units and SPE Metric Standards," Society of Petroleum Engineers, P.O. Box 833836, Richardson, TX 75083.

APPENDIX 1A

Definition of SI Base and Supplementary Units

	Unit	Symbol
Meter	m	The meter is the length equal to 1 650 763.73 wavelengths in vacuum of the radiation corresponding to the transition between the levels $2p_{10}$ and $5d_5$ of the krypton-86 atom.
Kilogram	kg	The kilogram is the unit of mass (and is the coherent SI unit); it is equal to the mass of the international prototype of the kilogram.
Second	s	The second is the duration of 9 192 631 770 periods of the radiation corresponding to the transition between the two hyperfine levels of the ground state of the cesium-133 atom.
Ampere	A	The ampere is that constant current which, if maintained in two straight parallel conductors of infinite length of negligible circular cross-section and placed one meter apart in vacuum, would produce between these conductors a force equal to 2×10^{-7} newton per meter of length.
Kelvin	K	The Kelvin, unit of thermodynamic temperature, is the fraction 1/273.16 of the thermodynamic temperature of the triple point of water.
Mole	mol	The mole is the amount of substance of a system which contains as many elementary entities as there are atoms in 0.012 kilogram of carbon-12. *Note* – When the mole is used, the elementary entities must be specified and may be atoms, molecules, ions, electrons, other particles, or specified groups of such particles.
Candela	cd	The candela is the luminous intensity, in a given direction, of a source that emits monochromatic radiation of frequency 540×10^{12} hertz (Hz) and that has a radiant intensity in that direction of 1/683 watt per steradian.
Radian	rad	The radian is the plane angle between two radii of a circle which cut off on the circumference an arc equal in length to the radius.
Steradian	sr	The steradian is the solid angle which, having its vertex at the center of a sphere, cuts off an area of the surface of the sphere equal to that of a square with sides of length equal to the radius of the sphere.

APPENDIX 1B

Definitions of SI Derived Units having Special Names

1. ***Absorbed dose*** - The gray is the absorbed dose when the energy per unit mass imparted to matter by ionizing radiation is one joule per kilogram.
   Note - The gray is also used for the ionizing radiation quantities: specific energy imparted, kerma, and absorbed dose index, which have the SI unit joule per kilogram.
2. ***Activity*** - The becquerel is the activity of a radionuclide decaying at the rate of one spontaneous nuclear transition per second.
3. ***Celsius temperature*** - The degree Celsius is equal to the Kelvin and is used in place of the Kelvin for expressing Celsius temperature (symbol t) defined by the equation $t = T - T_0$ where T is the thermodynamic temperature and $T_0 = 273.15$ K by definition.
4. ***Dose equivalent*** - The sievert is the dose equivalent when the absorbed dose of ionizing radiation multiplied by the dimensionless factors Q (quality factor) and N (product of any other multiplying factors) stipulated by the International Commission on Radiological Protection is one joule per kilogram.
5. ***Electric Capacitance*** - The farad is the capacitance of a capacitor between the plates of which there appears a difference of potential of one volt when it is charged by a quantity of electricity equal to one coulomb.
6. ***Electric conductance*** - The siemens is the electric conductance of a conductor in which a current of one ampere is produced by an electric potential difference of one volt.
7. ***Electric inductance*** - The henry is the inductance of a closed circuit in which an electromotive force of one volt is produced when the electric current in the circuit varies uniformly at a rate of one ampere per second.
8. ***Electric potential difference, electromotive force*** - The volt (unit of electric potential difference and electromotive force) is the difference of electric potential between two points of a conductor carrying a constant current of one ampere, when the power dissipated between these points is equal to one watt.
9. ***Electric resistance*** - The ohm is the electric resistance between two points of a conductor when a constant difference of potential of one volt, applied between these two points, produces in this conductor a current of one ampere, this conductor not being the source of any electromotive force.
10. ***Energy*** - The joule is the work done when the point of application of a force of one newton is displaced a distance of one meter in the direction of the force.
11. ***Force*** - The newton is that force which, when applied to a body having a mass of one kilogram, gives it an acceleration of one meter per second squared.
12. ***Frequency*** - The hertz is the frequency of a periodic phenomenon of which the period is one second.
13. ***Illuminance*** - The lux is the illuminance produced by a luminous flux of one lumen uniformly distributed over a surface of one square meter.
14. ***Luminous flux*** - The lumen is the luminous flux emitted in a solid angle of one steradian by a point source having a uniform intensity of one candela.
15. ***Magnetic flux*** - The weber is the magnetic flux which, linking a circuit of one turn, produces in it an electromotive force of one volt as it is reduced to zero at a uniform rate in one second.
16. ***Magnetic flux density, magnetic induction*** - The tesla is the magnetic flux density of one weber per square meter. In an alternative approach to defining the magnetic field quantities the tesla may also be defined as the magnetic flux density that produces on a one-meter length of wire carrying a current of one ampere, oriented normal to the flux density, a force of one newton, magnetic flux density being defined as an axial vector quantity such that the force exerted on an element of current is equal to the vector product of this element and the magnetic flux density.
17. ***Power*** - The watt is the power that represents a rate of energy transfer of one joule per second.
18. ***Pressure or stress*** - The pascal is the pressure or stress of one newton per square meter.
19. ***Electric charge, quantity of electricity*** - Electric charge is the time integral of electric current; its unit, the coulomb, is equal to one ampere second.

APPENDIX 1C

Summary of Basic Conversion Factors

LENGTH –
1 m = 3.281 ft = 39.37 in. = 100 cm = 1000 mm
1 ft = 0.305 m = 12 in. = 30.5 cm = 305 mm
1 statute mile = 1.61 km, 1 km = 0.621 statute mile

AREA –
1 m^2 = 10 000 cm^2 = 10.76 ft^2 = 1549 $in.^2$
1 hectare = 10 000 m^2 = 2.47 acres, 1 acre = 0.405 hectares
1 sq mile (section) = 2.59 km^2 = 259 hectares

VOLUME –
1 m^3 = 35.31 ft^3 = 6.29 bbl = 1000 L
1 L = 0.001 m^3 = 1000 cm^3 = 0.035 ft^3 = 61 $in.^3$
1 ft^3 = 0.0283 m^3 = 28.3 L
1 bbl (API) = 0.159 m^3 = 159 L = 5.61 ft^3

MASS –
1 kg = 2.205 lb = 1000 g
1 lb = 0.454 kg = 454 g
1 ton (metric) = 1000 kg = 2205 lb_m = 1 Mg

DENSITY –
1 kg/m^3 = 0.001 g/cm^3 = 0.0624 lb/ft^3
1 lb/ft^3 = 16.02 kg/m^3 = 0.016 02 g/cm^3
1 g/cm^3 = 1000 kg/m^3 = 62.4 lb/ft^3 = 1.0 kg/L

FORCE –
1 N = 0.225 lb_f = 0.102 kg_f
1 kg_f = 9.81 N = 2.205 lb_f
1 lb_f = 4.45 N = 0.454 kg_f

PRESSURE –
1 bar = 14.50 psi = 0.987 atm = 1.02 kg/cm^2
= 100 000 N/m^2 = 100 kPa (This text)
1 psi = 6.895 kPa

TEMPERATURE –
°C = 0.556 (°F – 32), K = °C + 273
°F = (1.8)(°C) + 32, °R = °F + 460

STANDARD CONDITIONS –

English			Metric		
Pressure	Temperature	Molal Volume	Pressure	Temperature	Molal Volume
14.696 psia	60°F	379.3 ft^3/lb-mol	100 kPa	15°C	23.96 m^3/kmol
14.65	60	380.5	100	0	22.71
14.73	60	378.4	101.325	15	23.64
14.50	60	384.4	101.325	0	22.41

APPENDIX 1C (continued)

METRIC UNITS –

Bar = 100 000 N/m^2 = 100 000 Pa = 1.02 kg/cm^2 = 100 kPa (This text)
Standard Conditions: 15°C and 100 kPa (This text)
1 000 000 std m^3 = 10^6 std m^3 = 41 740 kmol
There are 23.96 std m^3/kmol and 41 740 kmol/10^6 std m^3 (@ 15°C and 100 kPa)
Density of air at 15°C and 100 kPa = 1.21 kg/m^3
Density of gas at 15°C and 100 kPa = (γ_g)(1.21)
1 m^3 = 1000 liters; 1 liter = 1000 cm^3
1 g/cm^3 = 1000 kg/m^3

ENGLISH UNITS –

Standard Conditions: 60°F and 14.7 psia (This text)
1 000 000 std cu ft = 1 MMscf/d = 2636 lb–mol
Density of air at 60°F and 14.7 psia = 0.0764 lb/ft^3
1 ft^3 = 7.48 US gal = 6.23 UK gal
1 bbl = 42 US gal = 35 UK gal = 5.61 ft^3

CONVERSIONS BETWEEN UNITS –

1 std ft^3 (@ 60°F and 14.7 psia) = 0.0286 std m^3 (@ 15°C and 100 kPa)
1 std m^3 (@ 15°C and 100 kPa) = 34.92 std ft^3 (@ 14.7 psia and 60°F)
1 bbl = 159 L = 0.159 m^3
1 m^3 = 35.31 ft^3 liquid = 264 US gal = 220 UK gal = 6.29 bbl
1 L = 0.264 US gal = 0.22 UK gal = 0.0353 ft^3

Gas	Liquid
kmol/h = 1739 (10^6 std m^3/d)	kmol/h = 41.7 (m^3/d)(γ_L)/MW
kg/h = 50 372 (10^6 std m^3/d)(γ_g) = 1739 (10^6 std m^3/d)(MW of gas)	kg/h = 41.7 (m^3/d)(γ_L)
lb–mol/h = 110 (MMscf/d)	lb–mol/h = 14.6 (API bbl/d)(γ_L)/MW
lb/h = 3178 (MMscf/d)(γ_g) = 110 (MMscf/d)(MW of gas)	lb/h = 14.6 (API bbl/d)(γ_L)
Where: MW = molecular weight of gas or liquid	
Example: To get kmols per hour, multiply millions of meters cubed per day times 1739	

2

PLANNING THE SYSTEM

The gas conditioning and processing equipment is only a part of the entire system. The total system may look very much like that shown in Figure 2.1. For convenience, we divide each system into modules. A dehydration unit, for example, would be a module; as would a fractionation tower with its auxiliary equipment. The choice of module is governed by convenience, both for calculation and decision purposes.

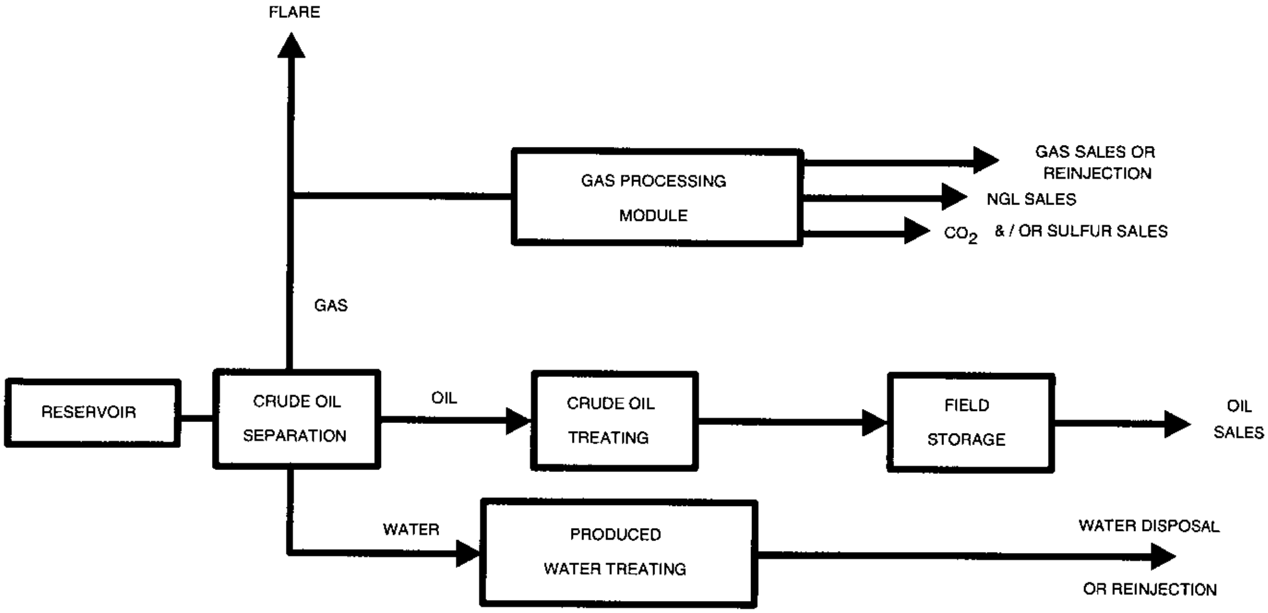


Figure 2.1 **Schematic View of a Total Production Processing System**

Unfortunately, one can do a sound job of designing, specifying and operating each modular unit and yet end up with a poor system. The reason ... each module has varying characteristics under varying loads that may result in a type of internal incompatibility. One modular unit may require a certain incoming analysis to produce the output desired. If a previous unit does not maintain this, subject unit may not prove satisfactory. The fault might not lie so much with that unit but with total system design (even though the unit is usually blamed).

Most of the errors I observe are what I call "errors of omission." Most facets of the problem receiving thoughtful, formal consideration usually are handled satisfactorily. It is the things we fail to consider properly that usually are at the root of most problems. One such omission is to concentrate on the detailed design of each module without proper consideration of the total system within which it resides. Another is failure to properly recognize the degree of uncertainty in the input and output specifications of the system--random variables within practical limits.

The process of simulation is nothing more than performing (in advance) those calculations which characterize system behavior. The most routine form of simulation simply involves solving the equations which (hopefully) describe the operation of concern. Although we currently do much of this on a computer, it adds nothing to the value of the result unless greater true precision is obtained. We may simply obtain more numbers in a given period of time. This in itself is good for more alternatives may be considered. But ... we much remember that better design is not an automatic advantage.

Total simulation must recognize formally the uncertainty (risk) of the numbers used. Using an average or most probable analysis is not an answer. These are only two points on the likely distribution curve (mean and mode respectively). Total simulation must include these concerns so that the system may possess necessary flexibility with minimum use of arbitrary safety factors.

Obviously, good samples are necessary. Appendix A considers sampling procedures necessary to obtain reliable data.

THE BASIC SYSTEM

Figure 2.1 represents a fairly complete processing setup for handling produced fluids. It encompasses almost all systems used. Not all elements shown are currently or potentially present in a given system. The purpose is to show most of the common alternatives. The time lag between original reservoir planning and the ultimate disposition of its "goods" (possibly many years hence) requires some initial concern for ultimate potential.

Each of the squares shown represents a *calculation module.* Within this module there is a body of equations and practice which enables one to design it – subject to the imposed constraints. Traditionally, adjectives have been used in front of the work "engineer" to loosely define the modular areas – chemical, petroleum, mechanical, etc. As the systems have become more complex, calculations within a module can never realistically exclude the other modules.

Not shown in the modular setup are the pumps, compressors, valves and fittings, and lines necessary to move, control and contain the fluids flowing between modules. These are inter-connecting modules difficult to show on diagrams.

Some major modules shown have a number of sub-modules representing component parts that involve some unique and/or separate engineering concern. For example, the NGL extraction module could be subdivided as shown in Figure 2.2. This figure is for the very simplest form of refrigeration system consisting of a wellstream exchange, refrigeration source, and separation of liquid from vapor.

The fact that all of the operations do not occur at, or in the vicinity of, the producing operation does not change the basic system or its needs. The very separation of the functions involved – resulting from organizational and geographical considerations – dictates the need for an overall planning function. Superimposing this necessary function on top of specific functions ... which have been at least semiautonomous ... is no easy task. On one hand, the planner does not always possess the technological expertise to impose realistic constraints on each individual element in the system. On the other hand, the people charged with operating each element resist change from those practices which have served them well – traditionally. Too often the charge seemingly reduces to, "Reduce cost," when what is meant is, "Increase profit."

Equally often, too much emphasis is placed on "people" and associated costs, within the same system, without adequate concern for the system itself. Profit will result from decreasing cost if all other factors remain the same. Unfortunately, overemphasis on cost usually changes other factors. Handling the system as

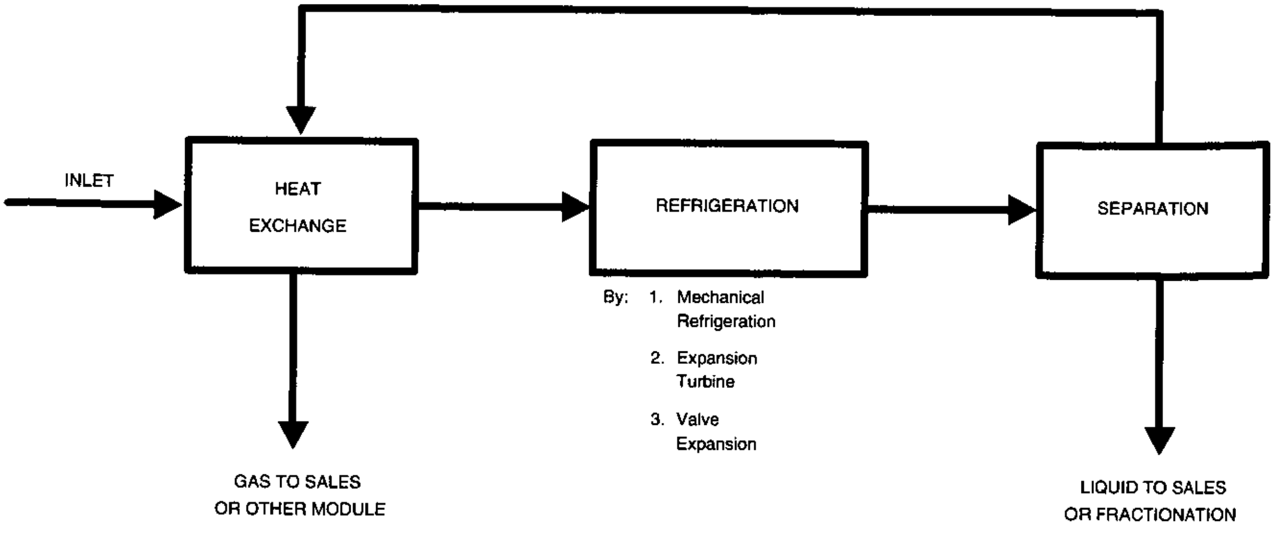


Figure 2.2 Refrigeration-Type Liquids Recovery Module

a system – instead of a series of loosely connected individual functions – can lead to a more rational basis for greater net profit.

Constraints of the Basic System

The system has several basic constraints:

1. The quantity and analysis of fluids entering
2. The market demand (quantity and price) for the effluent products
3. Legal and quasi-legal conditions imposed – "no-flare" gas orders, proration, contracts and agreements, national and political concerns, and the like
4. Environmental factors – labor availability and quality, climate, local customs, population density, availability of utilities and services, and the like
5. The risk tolerance level – technological, political and economic
6. The quantity and quality of available data

The problem of predicting the future is self-evident. The techniques for doing this must be reserved for another time. For our immediate purposes let us assume that a forecast of market and associated factors is available and that we possess some realistic measures of the uncertainties involved in that forecast.

At this point there are an infinite number of systems that could be devised to market the "stock" available for sale in the reservoir (theoretically). Actually, the choice is limited by a series of practical considerations.

The relative importance of each constraint varies with the individual system. No two systems are exactly alike even though they possess superficial similarity. Many of the errors made are by omission – we fail to recognize the sometimes subtle differences in the constraints.

As a practical matter, the first problem is developing the critical constraints (if any). Marketing is an obvious one but is outside the scope of this discussion. One comment, though, is pertinent – technological design must not only serve the present market efficiently but possess sufficient flexibility to accommodate a future market at minimum additional cost. For example, many reserves now exist in areas where there is no significant market for natural gas and natural gas liquids. However, any system design that is incompatible with future gas processing in these areas – without unnecessary additional cost – is a poor one.

Total reserves might be the paramount constraint. The maximum capital outlay that will yield a fair profit is fixed at some point by this concern. It is then only a matter of finding the best system at this, or hopefully lower, capital outlay to maximize profit.

An uncertain political climate might offer a similar constraint – limit the amount of risk capital to that which will afford both a realistic payout time (to reduce time risk) and a satisfactory rate of return.

These overall economic constraints provide the boundaries for our system "jigsaw puzzle." One then proceeds to the lower order, but equally important, legal and quasi-legal restrictions familiar to all. Compressor capacity rather than the reservoir may limit oil production when a "no flare" order is in effect. Fulfillment of a gas marketing contract may require a production schedule that is "inefficient" from the reservoir viewpoint alone. Many such restrictions are temporary but cause upset in the system.

THE DECISION MODULES

Several of the modules shown in Figure 2.1 are discussed to illustrate the considerations involved.

The Reservoir Module

A reservoir study generally is undertaken for one of two reasons – to establish value or to forecast performance under various production strategies, including enhanced recovery. The typical report deals in gross numbers not entirely suitable for production/processing planning and design. Needed is a special report showing greater detail about the character and condition of produced oil and/or gas.

Based on current samples, compositional balances can be made to forecast changes in gas and liquid analysis with time. These are very subjective but order of magnitude changes are detectable with sufficient accuracy to be of value in planning.

Geological data are valuable for judgment decisions involving the extrapolation of current data – number of wells, likelihood of solids production from core data, gathering system layout, etc.

On almost all new, large reservoirs some form of pressure maintenance is used to permit high initial production rates without excess pressure decline. The injection of water and/or gas usually is involved. At some point in time these will begin to "break through" into the production wells. Wellhead pressure will be different; liquid-gas ratios will change. Production/processing system needs will change accordingly. Is the surface system designed to accommodate only current conditions? If so, some major modifications will be necessary eventually. In an offshore or frontier environment the cost of modification plus the hidden cost of inefficient production practices can seriously compromise future profitability and limit reservoir recovery efficiency.

Any forecast of reservoir performance is inexact. But, experience has shown that the formal use of a proper forecast in planning and design considerations by qualified persons leads to more satisfactory surface systems.

The Separation Module

With few exceptions, some liquid will be obtained even though the fluid in the reservoir is all (or primarily) vapor, at reservoir conditions. In this instance, a flash calculation must be made at separation conditions to obtain the quantity and composition of all effluent streams.

If the primary effluent is crude oil or any other liquid stream, containing a reasonable percentage of heavy hydrocarbon molecules (larger than octane), this calculation is difficult. Gas specific gravity alone is inadequate for subsequent liquid recovery computations, less than adequate for even routine dehydration consideration.

Furthermore, even routine changes in temperature and pressure will affect the performance of subsequent modules.

Crude Oil Treating Module

This module is required to meet crude oil sales specifications:

1. BS&W (Basic Sediment and Water)
2. Vapor pressure
3. Salt
4. Sulfur content

The BS&W specifications is essentially an entrained water specification. It limits the amount of free water carried with the crude. It often varies from 0.3% to 3.0% by volume with the lower number applied to light crudes and the higher number to very heavy crudes (< 20°API). This specification is typically met by gravity separation. Heat, electricity chemicals, and mechanical coalescers may be used to enhance this process generally referred to as "crude oil dehydration." Chapter 11 covers the principles of this separation.

The vapor pressure specification limits the volatility of the crude oil. If the crude oil is stored or transported at or near atmospheric pressure this specification will often be equal to or less than 101.3 kPa [14.7 psia] at the system temperature. This specification can be stated in terms of a True Vapor Pressure (TVP) or a Reid Vapor Pressure (RVP). Chapters 5 and 16 present methods used to meet this specification.

The specification for salt and sulfur content are frequently met in the refinery rather than at the production facility. Salt is removed by mixing the crude with fresh water and removing the resultant brackish water in a crude oil dehydration module. Sulfur compounds may be removed by gas stripping chemical conversion or a combination of the two.

Produced Water Treating Module

Produced water must be treated in order to meet reinjection or disposal specifications:

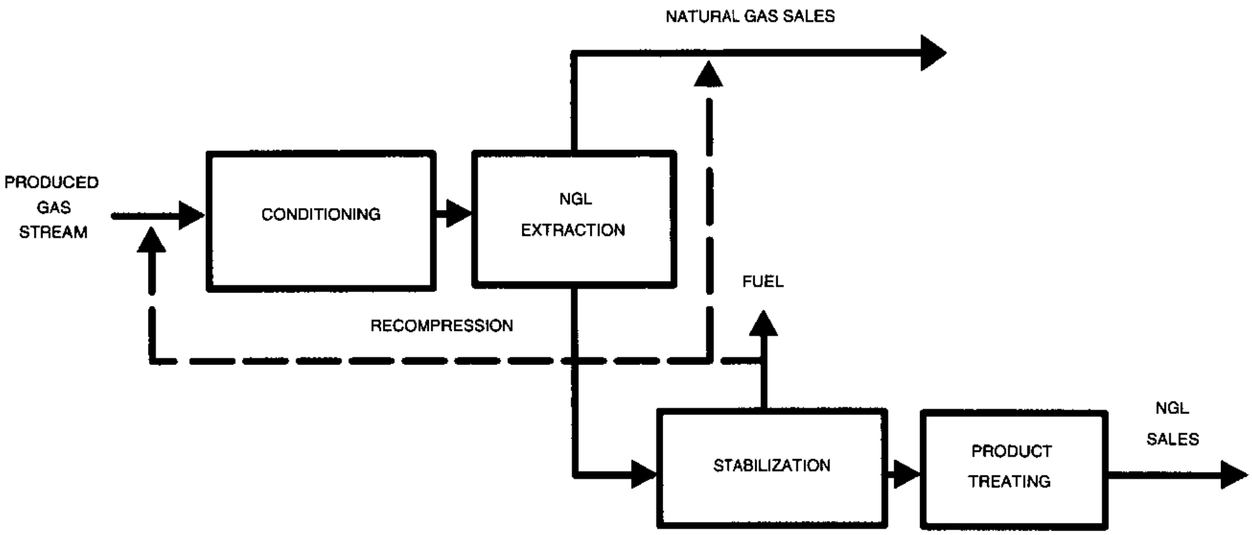
1. Hydrocarbons
2. Free solids
3. Dissolved solids e.g., $CaCO_3$, NaCl, $BaSO_4$, etc.

The hydrocarbon specification is particularly important if the produced water is discharged to the sea. For example, in the North Sea the oil content in the discharged water from an offshore platform is limited to 40 ppm by weight (monthly average). This specification is typically met by gravity separation, flotation units, centrifugal separation (hydrocyclones), or a combination thereof.

Free solids may require removal if the produced water is to be reinjected into the reservoir. Removal methods include gravity separation, filtration and centrifugal separation. Dissolved solids must be analyzed to assess their compatibility with connate water in the reinjection zone or with reinjection water from other sources such as sea water. In this case, specifications can only be established by detailed sampling and testing of the streams involved. These issues are discused in "Applied Water Technology," available from *Campbell Petroleum Series*.

Gas Processing Module

Figure 2.3 shows a simple flow diagram for a gas processing plant. This module may be used to condition the gas for sales, to extract and recover NGL's or both. The gas processing module may be further divided into sub-modules which are common to many gas processing facilities. These are discussed in further detail as follows:



Conditioning

1) Dehydration
2) Sweetening

NGL Extraction

1) Adsorption (SCU)
2) Absorption (lean oil)
3) Condensation
   a) Valve
   b) Turboexpander
   c) Refrigeration

Stabilization

1) Demethanizer
2) Deethanizer

Product Treating

1) Dehydration
2) Sweetening
   a) H_2S, COS, CS_2
   b) CO_2

Figure 2.3 Simple Flow Diagram for an NGL Extraction Plant

Gas Conditioning Module

This module typically exists at the inlet of the gas processing plant. It is generally designed to remove some or all of the following contaminants from the gas.

1. H_2O
2. Sulfur compounds H_2S, CS_2, COS, mercaptans, etc.
3. CO_2
4. N_2
5. O_2
6. Hg
7. Solids – hydrates, asphaltenes, dust, scale, etc.

Removal of water is termed gas dehydration. Several processes are used. The most common are:

Absorption
Adsorption
Condensation
Other– membranes, $CaCl_2$, etc.

In the absorption process, highly concentrated glycol solutions (triethylene glycol, TEG, is the most common) are used to physically absorb the water from the gas. This process is simple, effective and is probably the most widely used of the dehydration schemes. Glycol dehydration is discussed in Chapter 18.

The adsorption process utilizes a solid desiccant such as silica gel, alumina or molecular sieve (zeolite) to physically adsorb the water from the gas. It is typically used when very low (<1 ppm) effluent water contents are required and is discussed in Chapter 19

The condensation process is actually integrated with the refrigeration module for NGL extraction. Water condenses as the gas is cooled. In order to prevent hydrate formation, an inhibitor such as mono ethylene glycol (MEG) or methanol is injected into the gas stream. The principles of inhibition are covered in Chapter 6.

Membranes utilize a semipermeable polymer which allows certain molecules, e.g. H_2S, H_2O, CO_2, to permeate more rapidly than others, e.g. CH_4, C_2H_6, etc. The driving force for permeation is partial pressure. If the partial pressure of a component in the permeate stream is less than that in the feed stream the component will pass through the membrane. Selectivity depends on the membrane material. While the use of membranes specifically for the dehydration of natural gas is technically feasible, membrane applications to date have been primarily limited to CO_2 removal. Dehydration is a coincidental benefit of this process.

Calcium chloride, $CaCl_2$, can also be used to dehydrate natural gas. $CaCl_2$ will combine with water to form a brine which is removed from disposal. Use of this process is typically limited to small volume, remote applications.

Removal of H_2S and CO_2 from natural gas is covered in "Gas and Liquid Sweetening," Vol. 4 of the *Campbell Gas Conditioning and Processing series.* Several processes are available, many of them proprietary. In general, these processes can be summarized as follows:

Chemical absorption – amines, potassium carbonate
Physical absorption – SELEXOL, methanol
Solid bed – molecular sieve, iron sponge, zinc oxide
Other – direct conversion, membranes, extractive distillation

Chemical absorption using an amine based solvent is undoubtedly the most widely used of the above processes. It is used for both H_2S and CO_2 removal. Physical solvents may be more economical when H_2S and CO_2 concentrations are high. Solid bed processes are usually limited to H_2S removal and are competitive when concentrations are less than about 2000-3000 ppm. Direct conversion processes are applied for H_2S removal and convert the H_2S directly to elemental sulfur. Membranes and extractive distillation are applied for CO_2 removal.

Nitrogen removal from natural gas is achieved by cryogenic fractionation from methane. Operating temperatures vary with pressure but can be as low as –185°C [–300°F].

Mercury (Hg) removal from natural gas is necessary when aluminum plate fin exchangers are used in the gas processing module. Mercury removal is typically done by passing the gas across a bed of sulfur-impregnated activated charcoal where the mercury reacts to form mercuric sulfide, HgS. Effluent levels of Hg should be less than 1 ppb.

Natural Gas Liquids Extraction Module

Natural Gas Liquids (NGL) consist of hydrocarbon components in a produced gas stream which can be extracted and sold in their respective market. NGL products and their markets are summarized below:

Ethane (C_2) - Ethane has virtually one end use - petrochemical feedstock. It is the lightest NGL with a normal boiling point (NBP) of –88°C [–127°F]. Recovery can only be justified in those areas where a ready petrochemical market and viable transportation network exist.

Propane (C_3) - The market for propane is divided between petrochemical feedstock and fuel. Feedstock use is for light olefin manufacturing (ethylene, propylene). Fuel use is primarily residential and commercial in areas where other fuel sources (natural gas, heating oil, etc.) are not competitive. Its use as a transportation fuel is small, but growing. Often called LPG (Liquefied Petroleum Gas), propane is frequently sold as a mixture of propane and butane. Fuel demand tends to be cyclical, requiring large storage volumes (usually underground salt domes). Refrigerated storage (NBP = –42°C [–44°F]) is also used. For smaller volumes, above-ground storage under pressure is common.

Butanes (iC_4 and nC_4) - The market for butane is primarily petrochemical feedstock, fuel and gasoline blending. The end use for each isomer differs greatly. Isobutane (iC_4) is the more volatile isomer (NBP = –12°C [11°F]) and the most valuable. Its primary use is as a refinery feedstock for the manufacture of high octane blending components for gasoline. Normal butane (nC_4) is an important feedstock for the manufacture of monolefins (ethylene, propylene) and the diolefin, butadiene. Its NBP is –0.5°C [31°F]. When normal butane is used as a fuel it is normally blended with propane but can be used as a pure component. The largest use for butane is as a gasoline blending component for octane and vapor pressure control. As a result of their important gasoline and feedstock uses, butanes are normally fed to refineries.

Natural Gasoline (iC_5+) - Natural gasoline refers to the pentanes and heavier components in a gas stream. It consists primarily of straight and branched chain paraffins. Its most common use is refinery feedstock, although the petrochemical market for natural gasoline is growing.

NGL extraction is typically justified 1) to meet a gas sales specification requirement such as a hydrocarbon dewpoint or 2) to upgrade the market value of the produced gas and liquid streams. With the former, minimal extraction levels are required. Removal of the iC_5+ components is usually satisfactory. With the latter, the processor takes advantage of the price differential between a pure NGL product and its value as a natural gas constituent. An NGL product's value as a natural gas constituent is termed the *shrinkage* value. It is revenue foregone by the extraction of the NGL from the gas. This difference may be small, especially on the lighter NGL components, C_2 and C_3.

Economic justification of an NGL extraction facility requires that the market value of the NGL product be sufficient to cover all costs associated with its extraction. These costs include:

1. Shrinkage
2. Fuel
3. Maintenance and operating
4. Transportation and fractionation
5. Storage
6. Taxes, royalties, etc.
7. Capital

Commercial NGL extraction processes can be classified as follows:

Absorption – lean oil
Adsorption – Short cycle units, hydrocarbon recovery units
Condensation – mechanical refrigeration, turboexpander, valve expansion

The absorption process is frequently referred to as lean oil processing. NGL components are physically absorbed in a light hydrocarbon distillate stream (lean oil) and subsequently removed and recovered by stripping or distillation. The absorption process is frequently refrigerated to improve the efficiency. This was the dominant NGL extraction method up to the early 1970's. It has since been replaced on a widespread basis by the turboexpander process. This process is covered in more detail in Chapter 16.

The adsorption process is typically applied for hydrocarbon dewpoint control in special circumstances where other processes are not applicable or feasible. Recovery is limited to C_5+ components and H_2O. The simultaneous removal of hydrocarbons and water may be attractive in limited cases. This process is frequently referred to as a hydrocarbon recovery unit (HRU) or a short cycle unit (SCU). Additional information on hydrocarbon adsorption may be found in Chapter 19.

Condensation processes are the most widely used processes for the extraction of NGL from natural gas. Mechanical refrigeration plants utilize a commercial refrigerant such as propane or R-22 to chill the gas. Process temperatures are seldom less than about -40°C [-40°F]. This process is used both for hydrocarbon dewpoint control and NGL sales. Details of refrigeration are covered in Chapter 16.

Turboexpanders are widely used for NGL extraction The turboexpander is typically a single stage radial-inflow turbine. Expansion ratios vary from about 1.5 to 3.5 depending on the process objective. This process is employed for hydrocarbon dewpoint control as well as deep NGL recovery. Minimum process temperatures vary from 0° to -120°C [32° to -180°F]. Turboexpanders are discussed in Chapter 15.

Valve expansion is similar to turboexpansion except that the expansion device is a control valve rather than a turbine. This process is often referred to as an LTX, LTS or J-T (Joule-Thomson) process. It is sometimes used for hydrocarbon dewpoint control when high pressure feed gas is available from the reservoir. It can also be used for deep NGL recovery but this application is generally limited to small facilities. For similar expansion ratios, it is not as efficient as a turboexpander in cooling the gas. Chapter 16 provides a more detailed explanation of this process.

Stabilization Module

Once the NGL has been extracted from the gas, it must be stabilized to meet sales specifications. These specifications can vary significantly depending upon the end markets and transportation method. In some cases, the NGL product is a stabilized condensate consisting only of C_5+ having a vapor pressure less than 100 kPa [14.4 psia]. In other cases the NGL product is a C_4+ mixture which can be "spiked" into a crude oil stream for sale. In deep NGL recovery plants, the NGL product is more likely a C_2+ or C_3+ product which is transported to a fractionation plant for product separation, sale, and distribution.

The stabilization of the extracted NGL is typically accomplished by distillation. The distillation process may be refluxed or non-refluxed. Distillation is covered in Chapter 17. In some cases, particularly where the NGL product is a C_5+ stream with a vapor pressure less than 1 atm, flash stabilization can be used.

Product Treating Module

NGL must meet certain specifications regarding contaminant levels prior to sale. Contaminants of interest include CO_2, sulfur compounds, and water. In many cases these contaminants are removed from the gas in the conditioning module upstream of NGL extraction and are therefore not present in the NGL product. In some cases, however, these contaminants will be present in the NGL and must be removed. The most common applications of product treating involve the removal of

CO_2 from a C_2 or C_2+ NGL product
Sulfur compounds from LPG product

A typical CO_2 specification in ethane is 100 ppm. This is set by the petrochemical buyer. CO_2 is not easily fractionated from ethane since CO_2 and ethane form an aezeotrope. The most common method for removing CO_2 from ethane is absorption with an amine solution, typically DEA.

Sulfur compounds which may be present in LPG include H_2S, CO_S, CS_2, and mercaptans. These are typically present in low concentrations, less than 100 ppm, and are generally removed by adsorption on a molecular sieve.

Process Selection and Design Basis

As we have seen, several process alternatives are available for each module. Ultimately, the selection of the process depends on factors identified on page 27. Computer based process simulation programs are often used to provide preliminary numbers on which the process selection is based. A critical aspect of this feasibility study is design throughput.

Regardless of the process used, any simulation can proceed only from a knowledge of the quantity and composition of the entering stream. The problem is to find a size to meet all contractual requirements without excess capital investment. An oversized system, whose capacity is seldom utilized, results in excess capitalization charges. The problems of the undersized system are obvious. The real problem – how can one best accommodate the frequency and amplitude of the fluctuations as they are likely to occur? No mortal man can predict exactly when and to what degree a given fluctuation will occur. But ... with reasonable data, supplemented by reasonable judgment, one can predict the likely extreme conditions and the possible frequency of the intermediate conditions. Normal simulation, conditioned by the stochastic input, summarizes the procedure.

The quantity of throughput may be governed by the gas sale. This will vary. Individual wells and groups serving the module will suffer interruption. So now you have two constraints on both sides of the module – varying independently. Some combinations of these will defy prediction. Therefore, some confidence limits must be set up – the percentage of time that the plant will operate normally within the limits established during the design. Many use 95%. Assuming that one can keep the module going at less than desired efficiency most of the remaining time, true downtime is less than one percent.

Any confidence limit is achieved by built-in flexibility. A question arises how to achieve this. Once the limits are known, this is a mere mechanical design problem requiring normal expertise in the area.

CONTRACT TERMS

Gas and liquid contracts usually contain the following basic considerations:

Gas:

1. Minimum, maximum, and nominal delivery pressure
2. Maximum water content (expressed as a dewpoint at a given pressure or as a concentration)
3. Maximum condensable hydrocarbon content expressed as a hydrocarbon dewpoint, analysis, etc.
4. Maximum delivery temperature
5. Allowable concentration of contaminants such as H_2S, carbon disulfide, mercaptans, etc.
6. Minimum heating value
7. Cleanliness (allowable solids concentration)

Liquid:

1. Quality of products expressed as vapor pressure, relative or absolute density, or by standard designation such as Commercial Propane
2. Specifications such as color, concentration of contaminants, etc., as determined by standard tests
3. Maximum water content

Liquid products may be classified into two general categories – stock tank fluids from separators and fractionated products. The former is normally sold to a pipeline and is subject only to any pipeline limitations such as BS&W, specific gravity, vapor pressure, and presence of "light ends." It is sometimes referred to as a "slop" product to distinguish it from those products falling in the second category. In essence, the composition of this product will be fixed by the equilibrium relationships at the pressure and temperature of the storage tank.

In order to establish common terminology we will discuss some of the factors which affect product specification.

Gas Contracts

Satisfactory gas contracts are a primary requisite of a sound economic venture. Without them the degree of uncertainty introduced, with regard to revenue and source of gas, may make a potentially good prospect unattractive.

The following provisions are normally found in all gas contracts:

1. Definition of terms
2. Agreement to sell and purchase
3. Effective date and term of contract
4. Quantity and quality of gas
5. Delivery point and pressure
6. Operating stipulations
7. Priority rights of seller
8. Payment
9. Force majeure clauses
10. Title warranty of seller
11. Regulatory body provisions
12. Indemnity, default, and arbitration
13. Assignment of interest
14. Provisions for cancellation

Most of these items simply clarify the meaning, provide a mechanism for operating within the contract, or limit liability of the two parties involved. Many of these clauses represent legal verbiage insisted on by the legal profession. The effective date and the term of the contract are needed to establish the period involved.

Quantity and quality of gas. These, together with price, are the fundamental factors since they determine the processing equipment needed by buyer and seller, the total cost, and the revenue. Most contracts differentiate between casinghead and high-pressure, gas-well gas. A typical statement might be as follows:

> "The minimum average daily take by the buyer shall be one ten-thousandth (1/10 000) of the seller's reserves on all horizons classified by the state regulatory body as gas zones, except that the quantity so taken shall not exceed (some specified number). The minimum annual take shall be a quantity of gas equal to the minimum daily take times the number of days in the applicable year. The maximum take shall not exceed one and one-half times the minimum take. The buyer furthermore agrees to buy all casinghead gas produced from horizons classified as oil zones except that the guaranteed take shall not exceed (some specified number). In the event that the casinghead gas exceeds (some specified number) but is less than the total found by adding (some specified number) to the minimum gas well take, the buyer agrees to take the excess casinghead gas, provided that the gas well take is reduced accordingly. In no event will the buyer be required to purchase gas in excess of a sum obtained by adding (some specified number) to the minimum gas well take.

"It is recognized that on certain days the quantities of gas which buyer may desire to take hereunder may exceed the maximum take, and seller agrees, when and to the extent it may do so without damaging its wells or reservoir, by the use of its facilities, to deliver quantities of gas in excess of the maximum take, it being specifically understood, however, that the provisions of this paragraph shall not be construed as increasing the maximum take.

"Seller agrees to make available to buyer the maximum take each day, and buyer agrees to purchase and take during each contract year the minimum annual take."

Such a contract will normally set up a procedure to pay for any gas not taken annually under the agreement. The above provisions make this a take-or-pay contract. It assures the seller of a guaranteed minimum annual revenue, and the buyer has a reliable source of supply. It furthermore enables both parties to design their necessary facilities with maximum efficiency. The seller is protected from widely erratic buyer practices, and the buyer, in turn, is assured that the wells will be operated to assure a steady supply of gas.

The other extreme is the if, as, and when type of contract. In it the buyer often agrees to "take all gas," but then turns around and lists enough exceptions to invalidate the original statement. The disadvantages of this type of contract to the seller should be obvious.

Quality Provisions. Any or all of the following quality provisions are normally included in a sales contract:

1. **Heating Value**

   This may be expressed as a gross or net heating value per unit weight or volume of gas, or in therms. Where the gas contains noncombustibles like N_2 and CO_2, some factors or terms (like a Wobbe number) may be applied to limit this amount, in addition to heating value.

   The *gross heating value* is the heat produced on combustion of the gas with the theoretical amount of air required if the water formed by combustion is cooled to the reference temperature and then condensed. *Net heating value* is the comparable quantity when the water formed remains in the vapor state (the usual way it is done).

   The heating value may be expressed in many energy units. The kilojoule (kJ) or British Thermal Unit (Btu) are used most commonly. The following conversions are convenient.

   $$1 \text{ therm} = 100\,000 \text{ Btu} = 105.5 \text{ MJ}$$

   $$1000 \text{ Btu/ft}^3 = 37.2 \text{ MJ/m}^3 \, *$$

   *Standard conditions: 60°F and 14.7 psia, 15°C and 100 kPa

   Typical heating value specifications range from 35-45 MJ/m^3 [950-1250 Btu/scf] on a gross basis. These ranges may be significantly narrower for some end users.

2. **Sulfur Content**

   This limits the amount of sulfur compounds to prevent corrosion, toxicity and odor when the gas is burned. A typical specification for H_2S is 4 ppm [0.25 gr/100 scf]. Total sulfur specifications are considerably higher and vary with the buyer. It is usually expressed as equivalent H_2S.

   The specification of the allowable sulfur content in a "sweet" gas is often misunderstood. This gas is sweet only in the sense that there is no detectable, obnoxious odor on burning and it is not corrosive if the line is free of liquids. If the line contains liquids like water, glycol, amine and the like, this "sweet" gas can be corrosive. So ... in wet systems, meeting the specification is no assurance that corrosion will not occur.

3. **Maximum Temperature**

   A maximum delivery temperature at the transfer point may be specified. It is often about 49°C [120°F].

4. **Water Content (Dewpoint)**

   This specification will be stated as the mass of water per unit volume of gas or the maximum allowable water dewpoint temperature at a specified pressure. In the U.S. this specification is typically 7 lb H_2O/MMscf [110 mg/m^3]. This is equivalent to a water dewpoint of 32°F at 1000 psia. In Europe and Canada a water specification of 50-60 mg/m^3 [3-4 lb/MMscf] is common.

5. **Hydrocarbon Dewpoint**

   Today, most contracts specify that the gas shall be free of liquids, solids, dusts, gums and gum-forming constituents. Along with this, a specification is included that fixes the maximum allowable hydrocarbon dewpoint temperature at a given pressure. Hydrocarbon dewpoint specifications vary with climate and buyer. In North America and Europe typical values range from -10°C to 0°C [15-32°F] at any pressure.

6. **Other**

   Inerts (N_2, He, Ar) - Nitrogen is the most important inert. Its concentration is usually limited to less than 1-2%. Where produced concentrations are higher, seller usually negotiates higher allowances. N_2 content also affects heating value.

   CO_2 - Carbon dioxide is also an inert but is often specified separately because of its corrosive nature in the presence of water. A maximum value of 2% is typical, but the CO_2 specification is often negotiable.

7. The gas shall be produced in its natural state from wells or allowable processing facilities. If included, this means that the gas is conditioned only to meet the type of specifications summarized above; it is not processed to alter substantially the hydrocarbon content.

Delivery point and pressure. The delivery point is usually the valve on the meter run containing the buyer's meter. The contract then specifies the operation and testing of this meter run. The required delivery pressure is normally the maximum anticipated line pressure.

Priority rights of seller. In contracts between producer and processing plants the seller often is given the right to any residue gas needed for the development and operation of his lease, not to exceed his share of any such residue gas. If this quantity is not sufficient, the seller may furthermore reserve some of his gas for this purpose, exempt from contract provisions. Unless clarified, this provision may lead to difficulty since plant economics may depend on a reasonable residue gas sale. It is therefore important to stipulate the extent to which the lease operator may control the sale of residue gas.

Even where the plant operator feels he can take this gamble, the lending institution will not if gas revenue is essential in securing the loan. Consequently, the contract should specify whether or not such gas use is limited to heater or for drilling purposes, or also may be used in pressure maintenance or repressuring operations. Probably the best and cleanest contract specifies a certain minimum daily delivery. The operator may then do anything he pleases with any remaining gas (other than sell to someone else).

The seller may introduce air, gas or some extraneous substance into the well or formation when he feels that it is necessary for production. The buyer, of course, may then elect not to take the gas if such treatment renders it unmarketable, in his opinion. The seller may also clean out, deepen, or abandon any well without liability. In some instances the gas buyer has first option to purchase any abandoned well at salvage value.

General provisions. The force majeure clauses deal with the failure of either party to live up to the contract because of circumstances over which they have no control. This includes governmental rulings, acts of God, strikes, lockouts, riots, floods, storms, fires, explosions, or involuntary destruction of any sort. In the event of labor trouble, each party operates entirely within his own discretion. Notice of suspension due to force majeure must be given in writing.

The contract is normally binding on all heirs and assigns during its term, subject to existing governmental rules, and provides for the method of cancellation if the terms are not complied with. These are primarily legal rather than engineering matters.

Standard casinghead contract. Historically this has been a common type of contract used between plant and lease operators for casinghead gas. It is far from satisfactory, especially to many lease operators. In fact, many will sign one only as a last resort. Regardless of your position, it must be conceded that this contract has many drawbacks, because it contains many arbitrary and general provisions that leave room for different interpretations.

The primary complaint of operators is that the method of payment is complex and depends on inexact tests. Furthermore, the revenue is often negligible. From the plant standpoint, the disposition of the residue gas is not clearly specified. This points up the need for a contract to be as simple as possible, for people as a group are suspicious of things they do not understand. It must also be specific so that both parties are properly protected.

Because of these difficulties there has been a growing tendency to write specific contracts for given fields rather than to try to use a general form for all. Some clarification results by buying the gas outright for a fixed sum, this price reflecting average content, quantity, and residue price. Even though this has some drawbacks, it is more palatable to many producers, since it eliminates the testing and is simple. The price paid by the plant is usually 40% to 60% of the residue price and it retains title to all liquids.

Liquid Contracts

The liquid produced from gas contains many different specifications. Liquid produced by simple separation is a mixture whose composition is fixed by equilibrium as discussed in Chapter 5. This is sometimes called a "slop" cut.

All other liquid products are a result of a fractionation which separates a raw mixture into its component parts based on vapor pressure and other component physical properties. These are commonly called *natural gas liquids* and are produced from what are called NGL plants. If the effluent gas from an NGL plant is totally liquefied, it is called liquefied natural gas (LNG).

The amount of processing done at the production site depends on the amount of fluids, available transportation to market, and local conditions. Offshore, swamps, jungle or arctic type locations limit the feasibility of more complicated systems. The accent is on merely doing the least necessary at the site to transport the production to more favorable surroundings for future processing.

In an NGL installation one or more of the following products may be specified. The definitions are those of the Gas Processors Association.(2.1)

Natural gasoline. This is a mixed product whose basic specification is vapor pressure (as discussed in Chapter 4). As a general rule it must meet all of the following specifications.

Reid vapor pressure: 70-235 kPa [10-34 psia]
Percentage evaporated at 60°C [140°F]: 25-85%
Percentage evaporated at 135°C [275°F]: not less than 90%
End point: not higher than 190°C [375°F]
Corrosion: not corrosive by specified test
Doctor test: negative
Color: not less than plus 25 (Saybolt)
Water content: as specified

Commercial ethane. This primarily is a chemical feed stock for the manufacture of plastics and associated materials. The specifications vary but are usually rigorous for contaminants may affect processing. Limitations on CO_2 and C_1 are particularly important.

Usually, no more than 60-80% of the ethane may be recovered without adversely affecting the marketability of the effluent gas under heating value specifications.

Demethanized and mixes. An increasing number of pipelines are purchasing a demethanized product containing ethane, propane, butane, and natural gasoline. There are no standard specifications on such mixtures, although the methane content is often limited to 1-3% of the ethane. The purity tests of natural gasoline usually apply.

The contract usually calls for periodic analysis of the stream. This is then used to divide the total purchase into natural gasoline, butanes, propane and ethane. The applicable rate for each component times its quantity is then added to establish the total revenue.

Commercial propane. This defines a fluid composed of at least 95% propane and/or propylene, whose true gauge vapor pressure must not exceed 1.45 MPa(g) [210 psig] at 38°C [100°F], and which satisfies GPA tests for total sulfur, residue, dryness, and corrosive compounds. Purchasers may offer about 90% of the resale price. This resale price is not tied to any base and fluctuates widely, depending on season and geographical location.

Commercial butane. A product meeting this designation is composed predominantly of butanes and/or butylenes and has a true vapor pressure not greater than 480 kPa(g) [70 psig] at 38°C [100°F]. The temperature at which 95% by volume has evaporated shall not exceed 1.1°C [34°F] when corrected to a barometric pressure of 100 kPa. It must meet the same general tests as propane for contaminating substances.

Butane-propane mixtures (LPG). The standard definition of this mixture is that it shall not have a true vapor pressure higher than commercial propane at 38°C [100°F] and shall pass the 95% boiling point test for butane. It must further pass all butane purity tests. In most instances it is sold for domestic heating service or used for secondary recovery of oil. The composition, for heating use, is varied to assure volatility in the various seasons, but the vapor pressure of the commercial product seldom exceeds 860 kPa(g) [125 psig] at 38°C [100°F].

Propane HD-5. This is a special grade of propane for motor fuel and other use which shall not have a true vapor pressure of more than 1.38 MPa(g) [200 psig] at 38°C [100°F]. It conforms to all other Commercial Propane tests except that it cannot contain over five liquid volume percent propylene and must contain at least ninety liquid volume percent propane.

It is usually easier to obtain a take-or-pay contract on LPG than commercial propane or butane, particularly if it may be marketed locally or the plant is near underground storage. Many small plants are able to find a suitable local outlet. In any event, the price policy is about the same as for propane. Very large quantities will, of course, command special consideration contract-wise.

Traditionally, in some markets, NGL liquids have been "spiked" back into crude oil to upgrade crude oil quality and price. A more modern approach is to use crude oil stabilization instead of gas-oil separation. It is more efficient in many situations. The stabilizer retains the maximum amount of NGL components in the crude that the liquid sales contract permits.

THE PROJECT PLAN

The most critical phase of project development occurs at or near the very beginning. At this point a series of philosophical guidelines must be applied to all calculations.

If the input is from a petroleum reservoir, one must recognize that future performance is not absolutely controllable. The reservoir engineering does not dictate how said reservoir will perform in the future; current analyses and flow conditions will certainly change with time. A whole series of questions like those following must be answered.

> What is the probable forecast of reservoir performance? What is the trend, magnitude and timing of fluid analyses, gas-oil ratios, etc.?
>
> What are the magnitude and timing of enhanced recovery techniques?
>
> Based on the above, what modifications of current analytical data are desirable?
>
> Are current data from tests, samples and analyses indicative of at least current reservoir characteristics? Do such data reflect equilibrium reservoir performance? If production is to be from more than one zone, have all such zones been tested and sampled?
>
> Is the reservoir likely to produce solids at the flow rates anticipated?
>
> What amount of water must be handled, now and in the future?
>
> If water and/or gas is being injected, when and how will breakthrough occur? Is a lift system anticipated? What will the impact be on the surface system?

None of the answers to questions like these will be absolute. They may be classified as very likely, probable and possible.

Very likely:	almost certain to occur; exact timing and magnitude uncertain but suitable estimates are possible. Need to include provisions for this occurrence in initial design.
Probable:	not so certain to occur; odds are 50-50 or slightly better. System should be designed so that it can be modified efficiently to handle this occurrence if it indeed happens. Some additional, initial capital cost is justified to accomplish this.
Possible:	could occur but the likelihood is very low. Warrants some consideration but no additional capital outlay – may change layout of system to accommodate future modification.

If the above type of exercise is undertaken by a qualified team of professionals who communicate effectively, the end result will be superior to most existing systems.

The same exercise is desirable at the output end of the system – determining what products to produce. This is a repeat of the above. The product market can be very volatile. Forecasts of prices and demand can be more inaccurate than reservoir predictions. In the 1970's, for example, one would seriously have considered recovery of liquid ethane from natural gas. The market and the price for ethane were excellent and growing. In the early 1980's, ethane recovery did not appear attractive in some areas. The market demand was flat and price was not attractive; it was more profitable to sell ethane in the natural gas.

There are cases where natural gas found in remote areas has no value through normal sale as gas. The cost f.o.b. the market place is (and will continue to be) higher than the market will bear. It possesses no current economic value but needs to be saved for future generations. The strategy evolved must reflect this reality.

Once the input and output have been analyzed properly, selection of the system to get between the two can then begin. The first step needed is a *conceptual design*; what often happens is a mere *feasibility study*. The latter, as the word "feasibility" implies, may be nothing more than how to modify a preferred process scheme to address the problem at hand. Licensing is a potential limitation. Have you ever heard an engineer-contractor recommend a process licensed to a competitor? No!

A true conceptual design looks at all processes, selects the best one and then makes a preliminary optimization of the one chosen. A process flow drawing (PFD) is included in the report. This is then reviewed as a preliminary to preparation of final design specifications.

It is very important in this review to consider the effect of modules on each other. As the input conditions to the first module vary, so must the output to some degree. The input to the second module varies, and so on down the line. This interdependence of modules can lead to a bad system even though no technical mistakes were made on the design of individual modules. In the final analysis, each module must be designed with enough flexibility to remain compatible with the others as system conditions change.

Failure to do this early planning, as described generally above, is the major cause of poor systems and costs the industry untold millions of dollars in profit. If this phase is done properly, the conversion of ideas into hardware is relatively routine.

SUMMARY

The philosophical discussions in these first two chapters have been included to emphasize that engineering is a philosophical science whose proper application depends to a large degree on human judgment. Since we can now out-compute our real knowledge, the calculation itself is no longer the primary challenge.

In spite of this, correct and meaningful calculations are still important. As we discuss the principles, laws, correlations, and equations that have proven useful, please remember: calculations solve equations and models, but not problems. Only people solve problems.

REFERENCES

2.1 Gas Processors Association (GPA), Tulsa, Oklahoma.

NOTES:

3

PHYSICAL PROPERTIES OF HYDROCARBON SYSTEMS

A reliable estimate of physical properties is required in order to obtain reliable calculations. For single component systems data are readily available. Tables 3.1 and 3.2 are examples of such data that will be used for calculations. References 3.1-3.3 may be used to estimate properties of pure paraffins heavier than decane.

Note: Table 3.2 is in two parts. Table 3.2(a) is in SI metric units and Table 3.2(b) is in English units. This is the system used throughout this book when corresponding tables and figures are shown in separate units.

TABLE 3.1

Physical Constants

[illegible]

Compound	Molecular Weight	z_c	Critical Temperature °R	Critical Temperature K	Critical Pressure psia	Critical Pressure MPa	EMR	Acentric Factor ω
C_1	16.043	0.29	343	191	666	4.60	14.19	0.0104
C_2	30.070	0.29	550	305	707	4.88	24.37	0.0979
C_3	44.097	0.28	666	370	617	4.25	34.63	0.1522
iC_4	58.124	0.28	734	408	528	3.65	44.74	0.1852
nC_4	58.124	0.27	765	425	551	3.80	44.24	0.1995
iC_5	72.151	0.27	829	460	491	3.39	55.30	0.2280
nC_5	72.151	0.27	845	470	489	3.37	55.27	0.2514
nC_6	86.178	0.26	913	507	437	3.01	65.58	0.2994
nC_7	100.205	0.26	972	540	397	2.74	75.88	0.3494
nC_8	114.232	0.26	1024	569	361	2.49	86.19	0.3977
nC_9	128.259	0.25	1070	595	332	2.29	96.53	0.4445
nC_{10}	142.286	0.25	1112	618	305	2.10	106.86	0.4898
nC_{11}	156.302	0.24	1150	639	285	1.97	117.17	0.5350
nC_{12}	170.338	0.24	1185	658	264	1.82	127.50	0.5620
N_2	28.016	0.29	227	126	493	3.40	9.71	0.0372
CO_2	44.010	0.28	548	304	1071	7.38	14.44	0.2667
H_2S	34.076	0.28	672	373	1300	8.96	20.28	0.0948
O_2	32.000	0.29	278	155	731	5.04	8.69	0.0216
H_2	2.016	0.30	60	33	188	1.30	4.23	−0.2202
H_2O	18.015	0.235	1165	647	3199	22.06		0.3443

TABLE 3.2(a)

Properties of Paraffin Hydrocarbons$^{(3.4)}$

Component	Methane	Ethane	Propane	iso-Butane	*n*-Butane	iso-Pentane	*n*-Pentane	*n*-Hexane	*n*-Heptane	*n*-Octane	*n*-Nonane	*n*-Decane
Molecular Weight	16.043	30.070	44.097	58.124	58.124	72.151	72.151	86.178	100.205	114.232	128.259	142.286
Boiling Point @ 101.3250 kPa (abs), K	111.63	184.57	231.08	261.34	272.66	300.99	309.21	341.89	371.57	398.82	423.97	447.31
Freezing Point @ 101.3250 kPa (abs), K	90.68	90.35	85.47	113.55	134.79	113.25	143.42	177.83	182.57	216.39	219.66	243.51
Vapor Pressure @ 313.15 K, kPa (abs)	(35 000.)	(6000.)	1341.	528.	377.	151.3	115.66	37.28	12.34	4.143	1.40	0.4732
Density of Liquid @ 288.15 K & 101.3250 kPa (abs)												
Relative density (water = 1)	(0.3)	0.3581	0.5083	0.5637	0.5847	0.6250	0.6316	0.6644	0.6886	0.7073	0.7224	0.7346
Absolute density, kg/m^3 (in vacuum)	(300.)	357.8	507.8	563.2	584.2	624.4	631.0	663.8	688.0	706.7	721.7	733.9
Apparent density, kg/m^3 (in air)	(300.)	356.6	506.7	562.1	583.1	623.3	629.9	662.7	686.9	705.6	720.6	732.8
Density of Gas @ 288.15 K & 101.3250 kPa (abs)												
Relative density (air = 1), ideal gas	0.5539	1.0382	1.5225	2.0068	2.0068	2.4911	2.4911	2.9753	3.4596	3.9439	4.4282	4.9125
Kilogram per cubic metre, kg/m^3, ideal gas	0.6784	1.2718	1.8650	2.4582	2.4582	3.0516	3.0516	3.6443	4.2373	4.8309	5.4259	6.0168
Volume @ 288.15 K & 101.3250 kPa (abs)												
Liquid, cm^3/mol	(50.)	84.04	86.84	103.2	99.49	115.6	114.3	129.8	145.6	161.6	177.7	193.9
Ratio, gas/(liquid in vacuum)	(442.)	281.3	272.3	229.1	237.6	204.6	206.8	182.1	162.4	146.3	133.0	122.0
Critical Conditions												
Temperature, K	190.55	305.43	369.82	408.13	425.16	460.39	469.6	507.4	540.2	568.76	594.56	617.4
Pressure, kPa (abs)	4604.	4880.	4249.	3648.	3797.	3381.	3369.	3012	2736.	2486.	2288.	2099.
Gross Calorific Value, Combustion @ 288.15 K & Constant Pressure												
Megajoule per kilogram, MJ/kg, liquid	–	51.586	50.008	49.044	49.158	48.579	48.667	48.344	48.104	47.919	47.783	47.670
Megajoule per kilogram, MJ/kg, ideal gas	55.563	51.920	50.387	49.396	49.540	48.931	49.041	48.722	48.482	48.290	48.137	48.043
Megajoule per cubic metre, MJ/m^3, ideal gas	37.694	66.032	93.972	121.426	121.779	149.319	149.654	177.556	205.431	233.286	261.189	289.066
Megajoule per cubic metre, MJ/m^3, liquid	–	18458.	25394.	27621.	28718.	30333.	30709.	32091.	33095	33865.	34485.	34985.
Volume air to burn one volume gas, ideal gas	9.54	16.70	23.86	31.02	31.02	38.18	38.18	45.34	52.50	59.65	66.81	73.97
Flammability Limits @ 310.93 K & 101.3250 kPa (abs)												
Lower, volume % in air	5.0	2.9	2.0	1.8	1.5	1.3	1.4	1.1	1.0	0.8	0.7	0.7
Upper, volume % in air	15.0	13.0	9.5	8.5	9.0	8.0	8.3	7.7	7.0	6.5	5.6	5.4
Heat of Vaporization @ 101.3250 kPa (abs)												
kJ/kg @ boiling point	509.86	489.36	425.73	366.40	385.26	342.20	357.22	334.81	316.33	301.26	288.82	276.06
Specific Heat @ 288.15 K & 101.3250 kPa (abs)												
C_p gas, kJ/(kg·K), ideal gas	2.204	1.706	1.625	1.616	1.652	1.600	1.622	1.613	1.606	1.601	1.598	1.595
C_v gas, kJ/(kg·K), ideal gas	1.686	1.429	1.436	1.473	1.509	1.485	1.507	1.517	1.523	1.528	1.533	1.537
$K = C_p/C_v$, ideal gas	1.307	1.194	1.132	1.097	1.095	1.077	1.076	1.063	1.054	1.048	1.042	1.038
C_p liquid, kJ/(kg·K)	–	3.807	2.476	2.366	2.366	2.239	2.292	2.231	2.209	2.191	2.184	2.179

TABLE 3.2(b)

Properties of Paraffin Hydrocarbons(3.4)

Component	Methane	Ethane	Propane	iso-Butane	*n*-Butane	iso-Pentane	*n*-Pentane	*n*-Hexane	*n*-Heptane	*n*-Octane	*n*-Nonane	*n*-Decane
Molecular Weight	16.043	30.070	44.097	58.124	58.124	72.151	72.151	86.178	100.205	114.232	128.259	142.286
Boiling Point @ 14.696 psia, °F	-258.73	-127.49	-43.75	10.78	31.08	82.12	96.92	155.72	209.16	258.21	303.47	345.48
Freezing Point @ 14.696 psia, °F	-296.44	-297.04	-305.73	-255.28	-217.05	-255.82	-201.51	-139.58	-131.05	-70.18	-64.28	-21.36
Vapor Pressure @ 100°F, psia	(5000.)	(800.)	188.4	72.58	51.71	20.445	15.574	4.960	1.620	0.5369	0.1795	0.0609
Density of Liquid @ 60°F & 14.696 psia												
Relative density @ 60°F/60°F	(0.3)	0.3562	0.5070	0.5629	0.5840	0.6247	0.6311	0.6638	0.6882	0.7070	0.7219	0.7342
°API	(340.)	265.6	147.3	119.8	110.7	95.1	92.7	81.60	74.08	68.64	64.51	61.23
Absolute density, lbm/gal (in vacuum)	(2.5)	2.970	4.227	4.693	4.870	5.208	5.262	5.534	5.738	5.894	6.018	6.121
Apparent density, lbm/gal (in air)	(2.5)	2.960	4.217	4.683	4.861	5.198	5.252	5.524	5.729	5.885	6.008	6.112
Density of Gas @ 60°F & 14.696 psia												
Relative density (air = 1), ideal gas	0.5539	1.0382	1.5225	2.0068	2.0068	2.4911	2.4911	2.9755	3.4598	3.9441	4.4284	4.9127
lb/M ft^3, ideal gas	42.28	79.24	116.20	153.16	153.16	190.13	190.13	227.09	264.06	301.02	337.98	374.95
Volume @ 60°F & 14.696 psia												
Liquid, gal/lb-mol	(6.4)	10.13	10.43	12.39	11.94	13.85	13.72	15.57	17.46	19.38	21.31	23.45
ft^3 gas/gal liquid, ideal gas	(59.1)	37.48	36.375	30.64	31.79	27.39	27.67	24.37	21.73	19.58	17.81	16.33
Ratio, gas/(liquid in vacuum)	(442.)	280.4	272.1	229.2	237.8	204.9	207.0	182.3	162.6	146.5	133.2	122.2
Critical Conditions												
Temperature, °F	-116.67	89.92	206.06	274.46	305.62	369.10	385.8	453.6	512.7	564.22	610.68	652.0
Pressure, psia	666.4	706.5	616.0	527.9	550.6	490.4	488.6	436.9	396.8	360.7	331.8	305.2
Gross Calorific Value, Combustion @ 60°F												
Btu/lb, liquid	–	22181.	21489.	21079.	21136.	20891.	20923.	20783.	20679.	20601.	20543.	20494.
Btu/lb, gas	23891.	22332.	21653.	21231.	21299.	21043.	21085.	20942.	20838.	20759.	20700.	20651.
Btu/ft^3, ideal gas	1016.0	1769.6	2516.1	3251.9	3262.3	4000.9	4008.9	4755.9	5502.5	6248.9	6996.5	7742.9
Btu/gal, liquid	–	65869.	90830.	98917.	102911.	108805.	110091.	115021.	118648.	121422.	123634.	125448.
Volume air to burn one volume, ideal gas	9.54	16.71	23.87	31.03	31.03	38.19	38.19	45.35	52.52	59.68	66.84	74.00
Flammability Limits @ 100°F & 14.696 psia												
Lower, volume % in air	5.0	2.9	2.0	1.8	1.5	1.3	1.4	1.1	1.0	0.8	0.7	0.7
Upper, volume % in air	15.0	13.0	9.5	8.5	9.0	8.0	8.3	7.7	7.0	6.5	5.6	5.4
Heat of Vaporization @ 14.696 psia												
Btu/lb @ boiling point	219.45	211.14	183.01	157.23	165.93	147.12	153.57	143.94	136.00	129.52	124.36	119.65
Specific Heat @ 60°F & 14.696 psia												
C_p gas, Btu/(lb-°F), ideal gas	0.5267	0.4078	0.3885	0.3867	0.3950	0.3844	0.3882	0.3863	0.3845	0.3833	0.3825	0.3818
C_v gas, Btu/(lb-°F), ideal gas	0.4029	0.3418	0.3435	0.3525	0.3608	0.3569	0.3607	0.3633	0.3647	0.3659	0.3670	0.3678
$K = C_p/C_v$, ideal gas	1.307	1.193	1.131	1.097	1.095	1.077	1.076	1.064	1.054	1.048	1.042	1.038
C_p liquid, Btu/(lb-°F)	–	0.9723	0.6200	0.5707	0.5727	0.5333	0.5436	0.5333	0.5280	0.5241	0.5224	0.5210

Most hydrocarbon streams are mixtures of the hydrocarbon species summarized in Chapter 1. Most contain some water and many contain varying quantities of contaminants like sulfur compounds (primarily hydrogen sulfide) and carbon dioxide. The physical and chemical properties depend on pressure, temperature, and composition.

The basic philosophy of the correlations depends on some understanding of the molecular character of the mixture. The primary factors involved are pressure, temperature, size and shape of individual molecules, their distance apart, the degree and nature of their attractive - repulsive forces, and the relative amount of each species present. These are not all independent variables so all correlations for mixtures are approximations of varying accuracy.

Each of the homologous series of hydrocarbons exhibit similar behavior through molecules with no more than about six carbon atoms. Shape and mass begin to have a greater effect on behavior as the molecules become larger. As a general rule, correlations for light hydrocarbons do not apply accurately for mixtures containing a significant quantity of "heavy ends."

Always remember that pressure and temperature are surrogate macroscopic measurements that reflect the effective microscopic energy of the individual molecules in the container. Pressure is the force exerted on the container wall, per unit area, by the kinetic energy of the molecules hitting said wall. The higher the kinetic energy the higher the pressure.

Increasing temperature adds energy to the molecules much of which often translates into kinetic energy. Thus pressure normally increases with temperature in a closed container. Molecules are attracted to each other until they get very close together at which point some repulsive forces begin to act. These forces effect the kinetic energy and the phase. When gas molecules get close enough together they possess the macroscopic characteristics of what we call a liquid. Further reduction of molecular energy leads to a solid. In a solid the molecular kinetic energy is relatively small and the molecules essentially vibrate in place.

As we will see later, change of phase requires a comparable change in the energy of the system.

The more similar the character of the mixture molecules, the more orderly their behavior. A single component system composed entirely of a simple, relatively spherical molecule like methane behaves in a very predictable, correlatable manner. There are not absolute rules of behavior but the accuracy of the calculation decreases as one goes down the following list:

- Single component system.
- Mixture of molecules from the same homologous series.
- Mixture of molecules from different homologous series.
- Hydrocarbon mixtures containing sulfur compounds and/or carbon dioxide.

Performance data for single component systems can be accurately correlated in graphical or tabular form. For all of the others, we normally must use some P-V-T equation of state along with what is usually called a "combination rule." The first is simply some equation containing the variables pressure, volume, and, temperature along with one or more empirical constants that are determined by comparing the equation with measured data for a given type of system. The breadth of reliability of the result is inexorably tied to the size and complexity of the data base.

In order to do this matching a given combination rule is used with the P-V-T equation. The principle of the combination rule is that the total mixture properties is some kind of summation of the properties of the molecular components of which it is composed. The simplest combination rule is to assume that the contribution of individual molecules is in proportion to their relative quantity in the mixture. This is never strictly true but serves as a reasonable approximation in many common applications.

The more dissimilar the molecules, the less accurate the prediction becomes. Sour fluids containing sulfur compounds and/or carbon dioxide are a disorderly lot. It depends on the application, but the reliability of this approach is seldom better than ±10% and has a high likelihood of being in the 15-20% range. The moral – don't fret about this just recognize it and make the system flexible enough to handle the uncertainty.

Some current computer solutions contain a series of constants which are attributed to some properties of the system. This may add greater precision in the calculation without a commensurate increase in accuracy. No matter how complex they may appear, all P-V-T equations of state are empirical. Their application beyond the data from which the constants were determined compromises the validity of the results. Also remember that when dealing with mixtures you must use the combination rule employed in determining the constants in the equation.

No mention of water has been made in the above, even though water in liquid or vapor form is present to some degree in all of the systems discussed herein. Liquid water is essentially immiscible in hydrocarbons. In the vapor phase it represents a trivial percentage of the total mass present (seldom more than one part per 1000, by weight). Because normal phase behavior calculations do not apply for water, we use special methods summarized in subsequent chapters.

Equations of state commonly use the values of P, V and T at the critical point, a specific point on the phase behavior curves shown in Chapter 4. Each molecular species has a unique critical point, as shown in Tables 3.1 and 3.2.

For each of the pure components shown in these tables, the critical values represent the maximum P and T at which a two-phase, vapor-liquid system can exist (as shown in Figures 4.1 and 4.2); above P_c and T_c only a single phase is possible. For mixtures, pseudocritical values will be calculated which are correlation constants only and are not a point on the phase curve (except by coincidence).

The correlations which follow are used commonly for hydrocarbon systems and are suitable for use in future calculations in this book. Volume 3 of this series(3.5) contains additional property information.

EQUATIONS OF STATE

Any equation correlating P,V and T is called an equation of state. It all started with the Boyle-Charles Laws which state that P times V is proportional to T. We can add an equal sign by adding a proportionality constant (R) – normally called a *universal gas constant.* The net result in practical form is

$$PV = nRT \tag{3.1}$$

Where:
P = absolute pressure
V = volume
n = number of mols of gas of volume V at P and T
T = absolute temperature

This is known as the *ideal gas concept.* If V is a specific volume (reciprocal of density), then n is the reciprocal of molecular weight.

For most gas mixtures Equation 3.1 is valid up to pressures of about 400 kPa [60 psi]. As pressure increases above this level, its accuracy becomes less and the system should be considered a nonideal gas.

P	V	T	R
kPa	m^3	K	8.314 (kPa)(m^3)/(kmol)(K)
MPa	m^3	K	0.008 31 (MPa)(m^3)/(kmol)(K)
bar	m^3	K	0.083 14 (bar)(m^3)/(kmol)(K)
psi	ft^3	°R	10.73 (psia)(ft^3)/(lb-mol)(°R)
lb/ft^2	ft^3	°R	1545 (psfa)(ft^3)/(lb-mol)(°R)

Nonideal P-V-T Equations

A large number of P-V-T equations have been developed to describe nonideal, real gas behavior. Each is empirical in that it correlates a specific set of data using one, or more, empirical constants. There can be as many such equations as there are people who correlate data. Some of the most common equations are summarized below. The inclusion, or exclusion, of an equation is not a reflection of the worth of an equation. It merely reflects degree of usage (social acceptability).

van der Waals

$$(P + a/v^2)\,(v - b) = RT \qquad (3.2)$$

Where: a and b are correlation constants, v is the molar volume

Benedict-Webb-Rubin (BWR)

$$P = RT\rho + (B_oRT - A_o - C_o/T^2)\,\rho^2 + (bRT - a)\,\rho^3 + a\alpha\rho^6 + (c\rho^3/T^2)(1 + \gamma\rho^2)\,e^{-\gamma\rho^2} \qquad (3.3)$$

Where: A_o, B_o, C_o, a, b, c, α, and γ are correlation constants; P is absolute pressure; T is absolute temperature and ρ is the molal density

Redlich-Kwong (RK)

$$P = \frac{RT}{v - b} - \frac{a}{T^{0.5}\,v\,(v + b)} \qquad (3.4)$$

Where: a and b are correlation constants, and v is the molar volume

Peng-Robinson (PR)

$$P = \frac{RT}{v - b} - \frac{a(T)}{v(v + b) + b(v - b)} \qquad (3.5)$$

The van der Waals equation is the "granddaddy" of P-V-T equations. It is seldom used today but serves as a basis for many concepts used in the correlation of mixture properties.

Equations 3.2, 3.4 and 3.5 are called *cubic* equations of state because they are cubic when written in terms of volume. The constants a and b are a function of the critical properties T_c and P_c and a third parameter, usually the acentric factor, ω. For mixtures, mixing rules are necessary to determine the values of a and b. Mixing rules will be discussed briefly in a later section.

A number of modifications to these equations of state have been published and are used in the oil and gas industry. The purpose of the modifications is to improve the validity of the model. Examples include the Soave modification to the RK equation (SRK) and the Starling modification to the BWR equation (BWRS).

The above equations are used primarily as a basis for computer solutions. They tend to offer more precision, but the real accuracy of the results may be no better than the simpler methods when one considers the quality of the P, T, and composition data available for use in predicting behavior of actual systems.

I must emphasize that no correlation is equally good for all gas mixtures. Like all empirical equations, equations of state can be extrapolated (beyond compositions on which they were determined) only with an inherent loss of accuracy. This accuracy is dependent to a large degree upon whether the gas is composed of similar molecules. If a natural gas is composed entirely of straight-chain, paraffin hydrocarbons, a simple correlation might suffice. If significant amounts of "different" molecules like sulfur compounds and carbon dioxide are present, a correlation based on "sweet" gases may not be reliable for this "sour" gas.

Corresponding States Concept

The *Corresponding states Concept* states that physical and thermodynamic properties, e.g. density, viscosity, vapor pressure, etc., which depend on intermolecular forces, are related to the critical properties in a universal way. Mathematically, another way of stating this concept is to say that if an equation of state for any fluid is written in terms of reduced properties (v/v_c, P/P_c, T/T_c) that equation is also valid for any other fluid.

Reduced properties are defined as:

$$T_r = \frac{T}{T_c} \qquad P_r = \frac{P}{P_c} \qquad v_r = \frac{v}{v_c} \tag{3.6}$$

where the subscript "r" denotes a reduced property and "c" a critical property. Critical properties for several natural gas components may be found from Table 3.1.

The Corresponding States Concept is only partially valid for real fluids, but nonetheless is a powerful correlation tool. As you would expect, it is most accurate for mixtures of similar molecules exhibiting little interaction. It is illustrated in Figure 3.1.

The vapor pressure curves for methane, and propane are shown in Figure 3.1(a). The terminus of each curve represents the critical point for that substance. Figure 3.1(b) shows the same curves plotted on reduced coordinates (T_r, P_r). Notice how the two reduced vapor pressure curves are almost identical. If the corresponding states theory were perfectly valid, these curves would be identical.

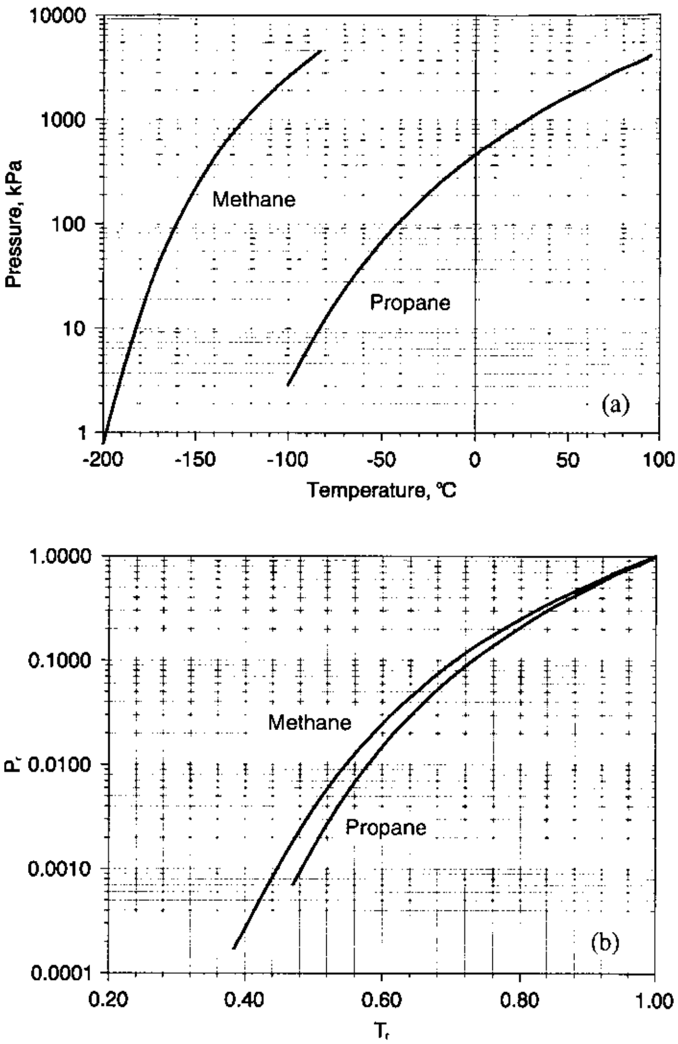


Figure 3.1 **Vapor Pressure (a) and Reduced Vapor Pressure (b) of Methane and Propane**

Third Parameters

The corresponding states concept says that fluid properties can be predicted from the relationship

$$\text{property, e.g. density,} = f(P_r, T_r) \tag{3.7}$$

As we have seen above, the corresponding state concept is not entirely valid. In order to improve the accuracy and generality of an equation of state a third parameter may be added to the corresponding states concept so that

$$\text{property} = f(P_r, T_r, \text{third parameter}) \tag{3.8}$$

where the third parameter is some term capable of characterizing fluid behavior.

A series of third parameters has been used in Equation 3.8. Three have been used most commonly.

Acentric Factor (ω). This was developed by Pitzer to describe the deviation of a fluid from simple fluid behavior, or its nonconformity with the corresponding states principle. For the simple, noble gases like argon, zenon, and the like, ω is equal to zero. It is close to zero for a gas like methane, as shown in Table 3.1.

The parameter ω is defined by the equation

$$\omega = -\log P_r - 1 \tag{3.9}$$

Where: $P_r = P^*/P_c$; P^* is the vapor pressure at $T = 0.7\ T_c$

This parameter is used widely as a correlating device. It is the third parameter used in the RK and PR equations of state. It can be visualized from Figure 3.1(b). It is a measure of the difference between the reduced vapor pressure of a component at $T_r = 0.7$ and the reduced vapor pressure of an ideal molecule at $T_r = 0.7$ which is $P_r = 0.1$.

Critical Compressibility (z_c). In Table 3.1 it may be noted that each component has a different z_c. This parameter has been used as a basis for z factor correlations designed for manual calculations. A chart of z versus P_r and T_r is prepared for different values of z_c. With the growth of computer usage this approach is not employed now as commonly as in the past, although it is a legitimate approach.

The parameters z_c and ω may be related using a correlation shown in Reference 3.4.

Molecular Refraction. This is a parameter related to the refractive index and molecular weight of a fluid. There are two forms – Lorentz-Lohrenz and Eykman. Both are used but the latter is more reliable for predicting gas properties.

The Eykman (EMR) possesses the advantage that values of EMR may be found readily from the molecular weight and specific gravity of the heaviest fraction shown in an analysis. This is usually heptanes plus (C_7+) in gas analyses.

Mixture Combination Rules

One cannot read a T_c and P_c for a mixture from a table. They must be calculated in some manner if P_r and T_r are to be used to calculate a property like z. Likewise, the values of the parameters a, b, c, ... etc. in Equations 3.2-3.5 must be determined in a special way for mixtures. Any equation, or series of equations, used to obtain mixture parameters is called a *combination rule* or *mixing rule.* The calculation, regardless of its exact form, is based on the premise that the properties of a mixture are some kind of weighted average summation of the properties of the individual molecules comprising that mixture.

The mixing rules used in cubic equations of state (i.e., van der Waals, Redlich-Kwong, Peng-Robinson) are

$$a_m = \sum\sum x_i\, x_j\, (a_i\, a_j)^{1/2}\, (1 - k_{ij}) \qquad b_m = \sum x_i\, b_i \tag{3.10}$$

Where: a_m and b_m = the a and b parameters for the mixture
a_i, a_j = a and b parameters for any components in the mixture
x_i, x_j = composition (mol fraction) for any two components in the mixture
k_{ij} = binary interaction parameter

Once a_m and b_m have been determined, the equation of state computations proceed as though a and b were for a pure component. With cubic equations of state the mixing rules sum the properties based on binary pairs. For example, a 4 component mixture would consist of 6 binary pairs.

The binary interaction parameter, k_{ij}, has no theoretical basis. It is empirical and is used to overcome deficiencies in the corresponding states theory or the basic model (equation of state). Binary interaction parameters are regressed from experimental data for a specific model and should be applied in that model only.

In addition, k_{ij}'s can be determined from regression of PVT data or VLE data. This will result in different k_{ij}'s for the same binary mixture.

A number of different mixing (or combination) rules are used in property correlations. Each property correlation is designed for use with the given combination rule used to prepare it from actual data. Like k_{ij}'s, these two go together and cannot be mixed. In fact, the correlation chart is prepared by regressing actual data using a specific mixture combination rule. The same is true for the multiple constant equations of state. The constants obtained apply only with the combination rule used to obtain them.

The critical values found from a combination rule are not necessarily true values; thus, they are called *pseudocriticals.* Remember: the word pseudo means "fake" or "not real." Thus, a pseudocritical value is not a point on the phase curve; it is merely a correlating parameter that uses the concept of corresponding states format. In this book a pseudocritical will be written with a "prime" – T_c' and P_c'.

GAS DENSITY

Because of its simplicity, Equation 3.1 is often corrected by a *compressibility factor*, z.

$$PV = zRT \tag{3.11}$$

for 1 mol or PV = nzRT for n moles.

Where: z is a function of equation of state parameters a, b, etc. in Eqns. 3.2-3.5 above.

Equation 3.11 may be written in the form

$$\rho = \frac{(P)(MW)}{zRT} \tag{3.12}$$

when the purpose is to calculate gas density at P and T. The units used for V and R dictate the units of ρ. It will be in g/cm^3, kg/m^3, or kg/L [lb/ft^3]; kg/m^3 is the preferred unit in most gas applications since in metric units most gas flows are shown in cubic meters.

In sections which follow, I will review some correlations for z which have proven useful for natural gas correlations.

Katz Correlation and Kay's Rule

This is the simplest method for z. It assumes that Equation 3.7 is valid. Figure 3.2 is the correlation prepared by Katz, *et al.*, for lean, sweet natural gas.

The values of P_c' and T_c' for use with Figure 3.2 may be found from Kay's Rule. This rule states that

$$P_c' = \sum y_i P_{c_i} \quad \text{and} \quad T_c' = \sum y_i T_{c_i} \tag{3.13}$$

Where: y_i is the mole fraction of each component in the mixture and T_{c_i} and P_{c_i} are the critical values for each component found from Tables 3.1 or 3.2

Dividing the actual, absolute temperature and pressure by the corresponding critical values gives the necessary reduced values.

Kay's Rule also may be used to calculate the effective molecular weight of a mixture, as shown in Example 3.1.

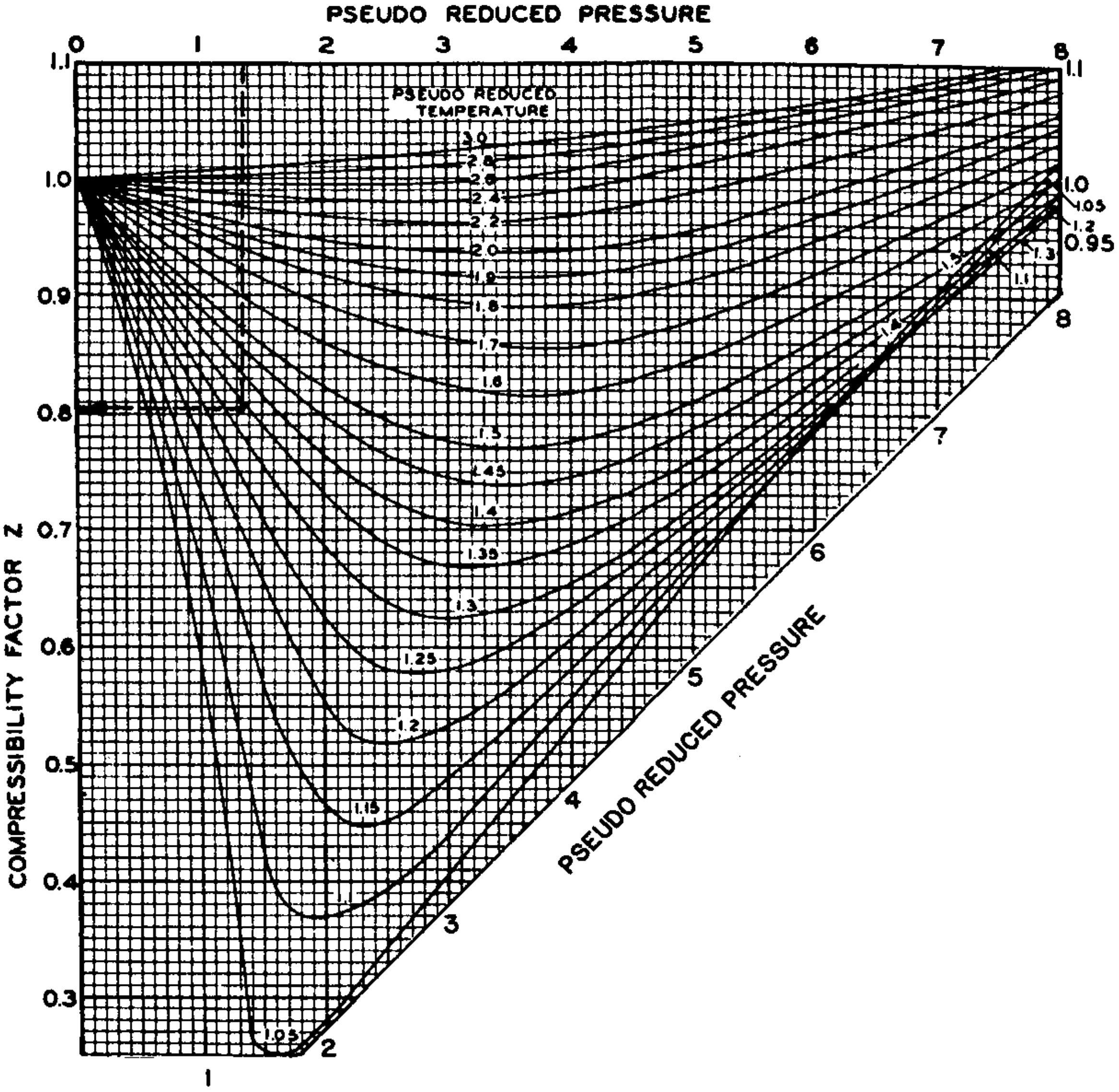


Figure 3.2 **Katz Compressibility Factor Chart**

Example 3.1: Calculate the compressibility factor for the gas shown in the table below at 13.94 MPa [2021 psia] and 331 K [595°R] using both Kay's Rule and Figure 3.2.

Comp.	y_i	English				Metric				ΣMW	
		P_c	P_c'	T_c	T_c'	P_c	P_c'	T_c	T_c'	MW	MW′
N_2	0.0046	493	2.3	227	1.00	3.40	0.02	126	0.6	28.016	0.13
CO_2	0.0030	1071	3.20	548	1.60	7.38	0.02	304	0.9	44.010	0.13
H_2S	0.1438	1300	186.9	672	96.60	8.96	1.29	373	53.6	34.076	4.90
C_1	0.8414	666	560.40	343	288.6	4.60	3.87	191	160.7	16.043	13.50
C_2	0.0059	707	4.20	550	3.2	4.88	0.023	305	1.8	30.070	0.18
C_3	0.0008	617	0.50	666	0.5	4.25	negl.	370	0.3	44.097	0.04
iC_4	0.0003	528	0.20	734	0.2	3.65	negl.	408	0.1	58.124	0.02
nC_4	0.0002	551	0.10	765	0.2	3.80	negl.	425	0.1	58.124	0.01
	1.0000		758 psia		392°R		5.23 MPa		218 K	MW = 18.90	

Example 3.1 (Cont'd.): **Metric:** $P_r' = 13.94/5.23 = 2.67$, $T_r' = 331/218 = 1.52$

English: $P/P_c' = P_r' = 2021/758 = 2.67$, $T/T_c' = T_r' = 595/392 = 1.52$

From Figure 3.2, Z = 0.80

Now, the actual density of the gas can be found from Equation 3.12.

$$\rho\ (\text{kg/m}^3) = \frac{(13.94)(18.90)}{(0.80)(0.008\ 31)(331)} = 120$$

$\frac{P M}{z R T}$

Application of Figure 3.2 to Gases Containing H_2S and CO_2

Since so many people use the Katz compressibility factor correlation, the question is often asked how it may be extended to gases containing H_2S and CO_2.

There are two methods available for this application.

1. The approach proposed by Robinson, *et al.*,(3.7)
2. The approach proposed by Wichert and Aziz(3.8)

Approach 1 is an earlier work. It was carefully done but was based on limited data, most of which were obtained on synthetic gas mixtures. For a number of reasons, we prefer the work of Approach 2 which is summarized below.

The approach utilizes an adjustment of the P_c' and T_c' which is found from Kay's combination rule, and which was used to prepare the Katz chart. It uses a correction parameter, ε, found from Figure 3.3.

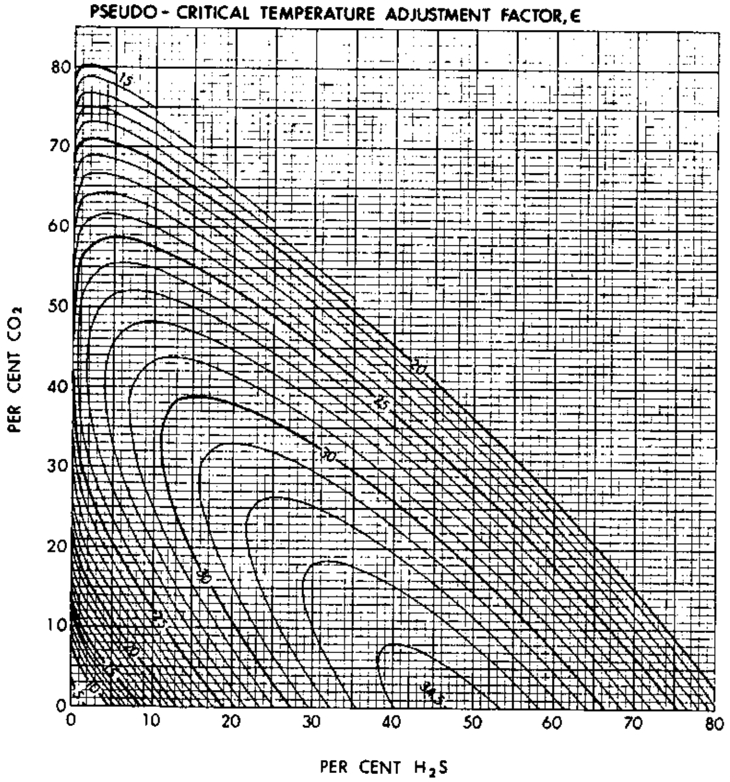


Figure 3.3 **Correction Factor Chart for Sour Gases(3.6)**

This correction parameter is used to adjust the pseudocriticals found from Kay's Rule. The adjusted values are then used to find the reduced pressure and temperature for use with Figure 3.2.

The equations for making the adjustment follow.

METRIC: $$T_c'' = T_c' - 0.556\,\varepsilon\ ,\quad P_c'' = \frac{P_c'\,T_c''}{T_c' + B(1-B)(\varepsilon)(0.556)} \tag{3.14}$$

ENGLISH: $$T_c'' = T_c' - \varepsilon\ ,\quad P_c'' = \frac{P_c'\,T_c''}{T_c' + B(1-B)\,\varepsilon} \tag{3.15}$$

Where: T_c'' and P_c'' = adjusted critical values for use in Figure 3.2
T_c' and P_c' = pseudocriticals from Kay's Rule, ε is from Figure 3.3
B = mol fraction H_2S in the gas

Figure 3.3 can be estimated by the equation[(3.9)]

$$\varepsilon = 120\,(A^{0.9} - A^{1.6}) + 15\,(B^{0.5} - B^4) \tag{3.16}$$

Where: ε = correction factor in °R (to obtain in K, divide by 1.8)
A = mol fraction H_2S plus CO_2 in the gas
B = mol fraction H_2S in the gas

Example 3.2: Rework Example 3.1 using Figure 3.3 correction.

From Figure 3.3, $\varepsilon = 21.0$

METRIC: $$T_c'' = 218 - (0.556)(21) = 206.3\ \text{K}$$

$$P_c'' = \frac{(5.23)(206.3)}{218 + (0.1438)(0.8562)(21.0)(0.556)} = 4.92\ \text{MPa}$$

ENGLISH: $$T_c'' = 392 - 21 = 371\text{°R}$$

$$P_c'' = \frac{(758)(371)}{392 + (0.1438)(0.8562)(21.0)} = 713\ \text{psi}$$

Then $P_r'' = 2.8$, and $T_r'' = 1.6$, and $Z = 0.83$

$$\rho\ (\text{kg/m}^3) = \frac{(13.94)(18.90)}{(0.83)(0.008\ 31)(331)} = 115$$

$$\rho\ (\text{lb/ft}^3) = \frac{(2021)(18.90)}{(0.83)(10.73)(595)} = 7.2$$

Approximate correlation for T_c' and P_c'

Often density calculations are required when very little information is available.

Figure 3.4 can be used to estimate pseudocritical temperature and pressure if no analysis is available. This is a very approximate correlation.

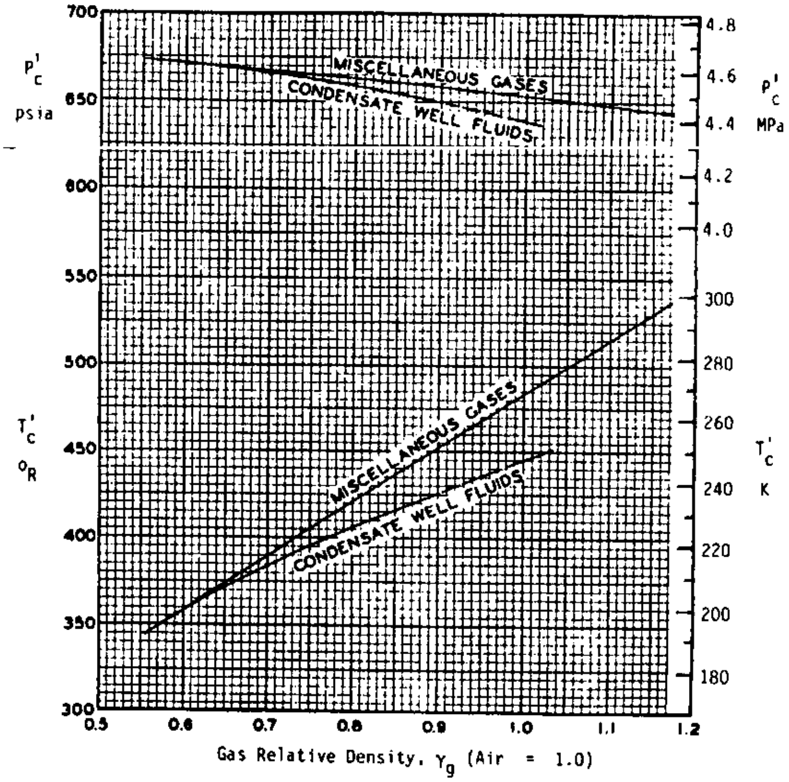


Figure 3.4 Approximate Pseudo-critical Properties of Natural Gas

Example 3.3: Rework Example 3.1 assuming only the molecular weight of the gas is known.

From the Example 3.1, MW = 18.90.

The relative density of this gas is 18.90/28.97 = 0.65

$P_c' = 4.62$ MPa [670 psia] and $T_c' = 208$ K [375 °R] From Figure 3.4

Metric: $P_R' = \dfrac{P}{P_c'} = \dfrac{13.95}{4.62} = 3.02 \qquad T_R' = \dfrac{T}{T_c'} = \dfrac{331}{208} = 1.59$

English: $P_R' = \dfrac{P}{P_c'} = \dfrac{2021}{670} = 3.02 \qquad T_R' = \dfrac{T}{T_c'} = \dfrac{595}{375} = 1.59$

From Figure 3.2, Z = 0.825

$$\rho\ (\text{kg/m}^3) = \frac{(13.94)(18.90)}{(0.825)(0.008\ 31)(331)} = 116$$

Other Correlations for z

Several correlations for determination of z can be found in the literature. Two correlations, EMR and SBV,(3.10, 3.11) which were included in previous editions of this text have proven useful.

In general, equation of state methods are probably the most widely used correlations for the determination of gas density (z). They are not necessarily the most accurate. Empirical correlations developed for a specific mixture or a narrow range of mixtures provide better accuracy but are less general. An example would be the Katz correlation which is quite good when applied to "sweet" pipeline quality gases, but less reliable for gases containing significant amounts of H_2S, CO_2, and/or N_2.

In order to extend the validity of empirical correlations, third parameters and more sophisticated mixing rules are required. This added complexity typically makes the calculation too tedious to perform manually – so a computer method is required.

For comparative purposes, the compressibility factor, z, has been calculated for the gas in Example 3.1 using several correlations. One should use the results as an indication of the range of results encountered, not as an indication of the inherent accuracy (or inaccuracy) of any one method.

	z
Katz	0.80
Katz w/ Wichert and Aziz correction	0.83
EMR	0.78
SBV	0.83
SRK equation of state	0.85
PR equation of state	0.81

LIQUID PHYSICAL PROPERTIES

Liquids are relatively incompressible compared to gases. However, the calculation of the properties of liquid mixtures is complicated by two factors – the presence of high vapor pressure components like methane and ethane, and the shrinkage in volume upon mixing two hydrocarbon liquids.

Both effects can be understood by remembering that liquid is made up of molecules of different sizes and shapes, possessing different amounts of energy.

The size and shape of the molecules affects all properties. If you magnified these molecules you would see a series of shapes and sizes with space between them. The size would depend on molecular weight. The larger paraffin molecules would be shaped somewhat like a sausage. The aromatic (ring) molecules would be more circular. The small molecules like methane and ethane would be rather spherical.

The energy of molecules in moving around is somewhat relative to vapor pressure, which in turn is related to molecular weight. Molecules like methane are very active, and this movement clears more space between molecules as collisions occur. The large molecules are more lethargic.

The properties are dependent on size, shape and energy. If you have a container full of large rocks, there are spaces in-between that can accommodate smaller-sized objects like gravel and sand. So ... if you mix liquids containing molecules of different sizes there is shrinkage on mixing as little molecules occupy voids between the bigger ones; i.e., the volumes are not additive.

All of this says that the physical properties of a mixture are dependent on things that are measures of energy, size and shape. Density is one of the basic factors. It is not only needed for basic calculations but correlates with things like viscosity, energy values, boiling points, etc.

Liquid correlations can be divided into two general categories: natural gas liquids (NGL) where an analysis showing most of the components is available, and undefined liquids (like crude oil) where a detailed analysis is not feasible.

If the NGL is recovered from the gas after any reservoir liquids have been separated previously, the amount of any component heavier than decane probably will be trivial and NGL correlations will suffice. If reservoir liquids are involved, some form of characterization is needed other than direct analysis. The distillation characteristics of the heavier components is the basic analytical tool employed.

When the liquid is composed of butane and heavier molecules their energy is low enough that relatively simple correlations may suffice. Some correction for pressure and temperature is required in many cases. Molecular energy increases with increasing temperature, thus reducing density and other physical properties. Increasing pressure has the opposite effect since it "squeezes" the molecules closer together.

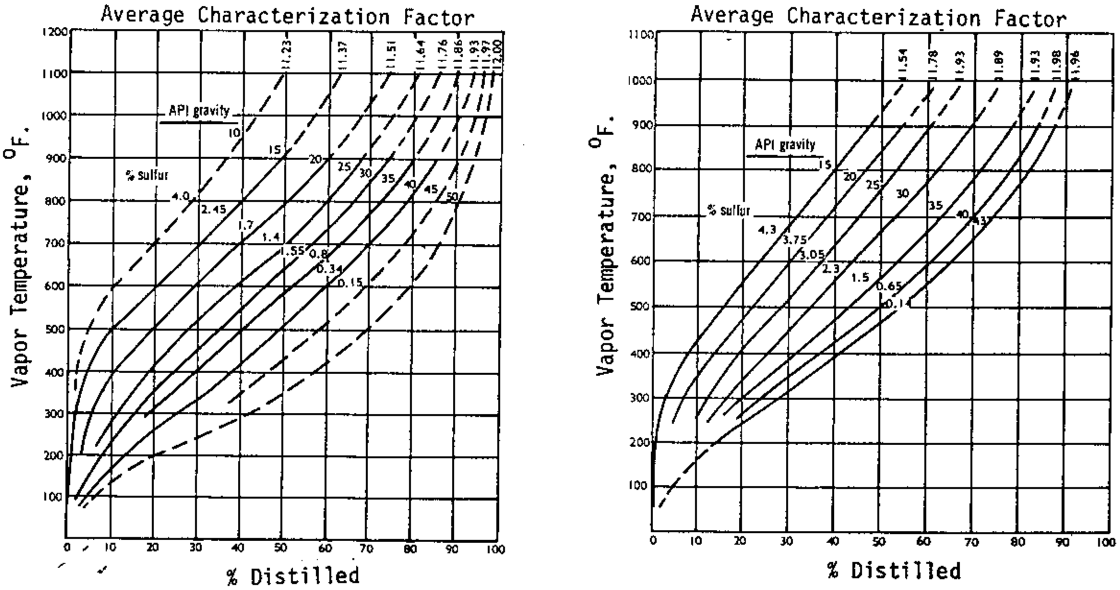
Distillation Characteristics

Two types of distillation are used commonly – ASTM and true boiling point (TBP). Both are batch distillations. The ASTM apparatus is a non-refluxed, single-stage unit. The temperature at which the first volume of liquid is collected overhead is recorded as the initial boiling point (IBP). The temperature is raised at a steady rate and recorded when several volume percentages of the sample have been recovered overhead. The end point temperature is recorded, usually when 95 volume % of the sample has been distilled overhead.

The TBP is the preferred approach. The apparatus usually contains ten or more equivalent stages. A high reflux rate is used to obtain a sharper separation between the subfractions distilled overhead.

The *average boiling point* often is used. This may be the 50% point of the distillation found from a combination rule or by a correlation. In addition to the correlations shown herein, further information is available from References 3.4, 3.6 and 3.12.

Figure 3.5 shows some representative TBP curves. The one on the left is based primarily on oil samples from North and South America. The one on the right is based on 85 samples from the Middle East and North Africa.



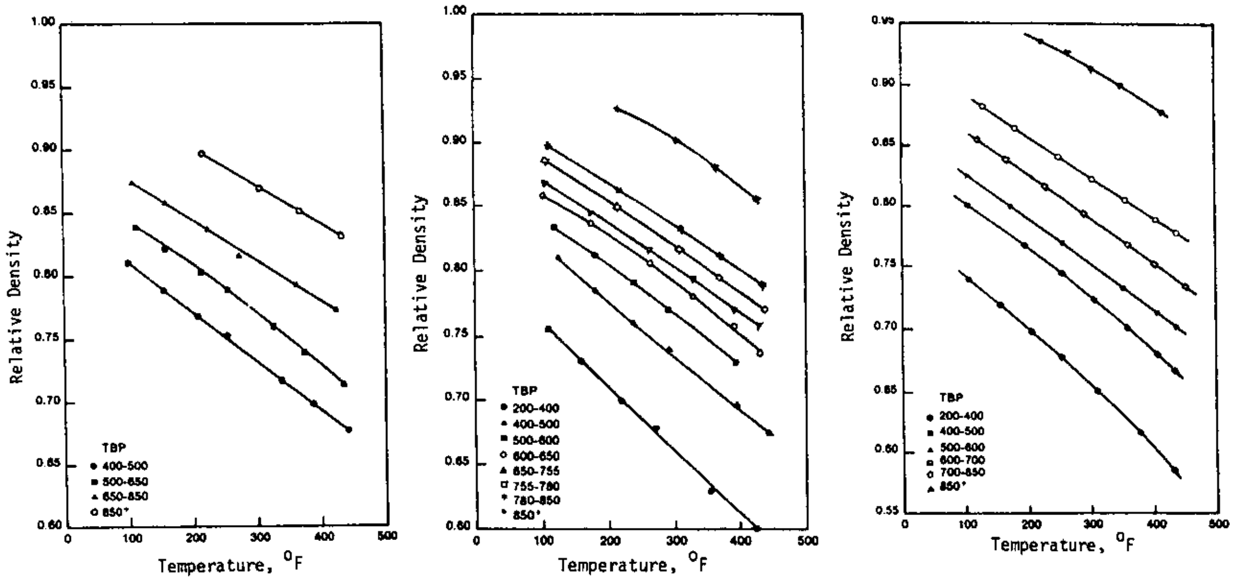
Average of About 550 Samples

Average Mid-East and N. Africa Samples

Figure 3.5 Some Representative TBP Distillation Curves

Sulfur affects the shape of TBP curves. Most sulfur occurs in the highest boiling or residual parts of the oil. Thus, the primary effect of sulfur is to steepen the slope of the higher boiling parts of the curve. Without sulfur the curves for the lower API gravity oils would be flatter.

As shown in Figure 3.6, density of a subfraction increases with TBP boiling range. The exact effect varies with each oil. Thus, data like that in Figures 3.5 and 3.6 are useful for illustration purposes only. One needs a sufficient quantity of a reliable sample for the specific liquid of concern to make definitive calculations.



North Sea Crude Fractions

U.S. Mid-Continent Crude Fractions

Arabian Liqht Crude Fractions

Figure 3.6 Liquid Density for TBP Distillation Cuts from Three Different Oils(3.13)

If the quantity of the liquid sample is sufficient, properties such as density and molecular weight of each subfraction may be determined. In turn, correlations are available to estimate the critical properties and acentric factor of each subfraction. For purposes of predicting physical and thermodynamic properties each subfraction may now be treated as an individual component. The subfraction actually consists of many different types of molecules (perhaps hundreds) and as such is referred to as a *pseudo-component*. Reference 3.4 provides an excellent summary of this procedure.

Watson Characterization Factor

This is for liquid

This factor is defined by the equation

$$K_w = \frac{A\,(T_b)^{1/3}}{\gamma} \tag{3.17}$$

		Metric	English
Where:	T_b = average boiling point	K	°R
	γ = relative density (sp. gr.)	–	–
	A = conversion factor	1.22	1.0

The Watson factor depends on composition. For paraffin base crudes, the value of K_w is about 11.9-12.2; 11.9 is normally used as a standard. The paraffin components all have values of K_w above 12.0; propane is 14.7; n-butane is 13.5, etc. The aromatics have values below 11.5. So ... the lower the value below 11.9 the larger is the amount of aromatics.

Figures 3.7 and 3.8 show the relationship between various fluid property parameters. The nomograph(3.14) is particularly useful. Knowing any two of the parameters enables one to estimate the others. A 40° API crude with a Watson Characterization factor of 11.9 will have an estimated molecular weight of 209 and an average boiling point of 255°C [491°F]. The carbon-to-hydrogen weight ratio varies with molecule size and type. It is another way to characterize the oil properties. Paraffin series molecules all possess a C-H weight ratio of less than 6.0. Thus, any value above 6.0 indicates other "families" are present. For most applications, the weight ratio can be ignored, the other four parameters shown being the major consideration.

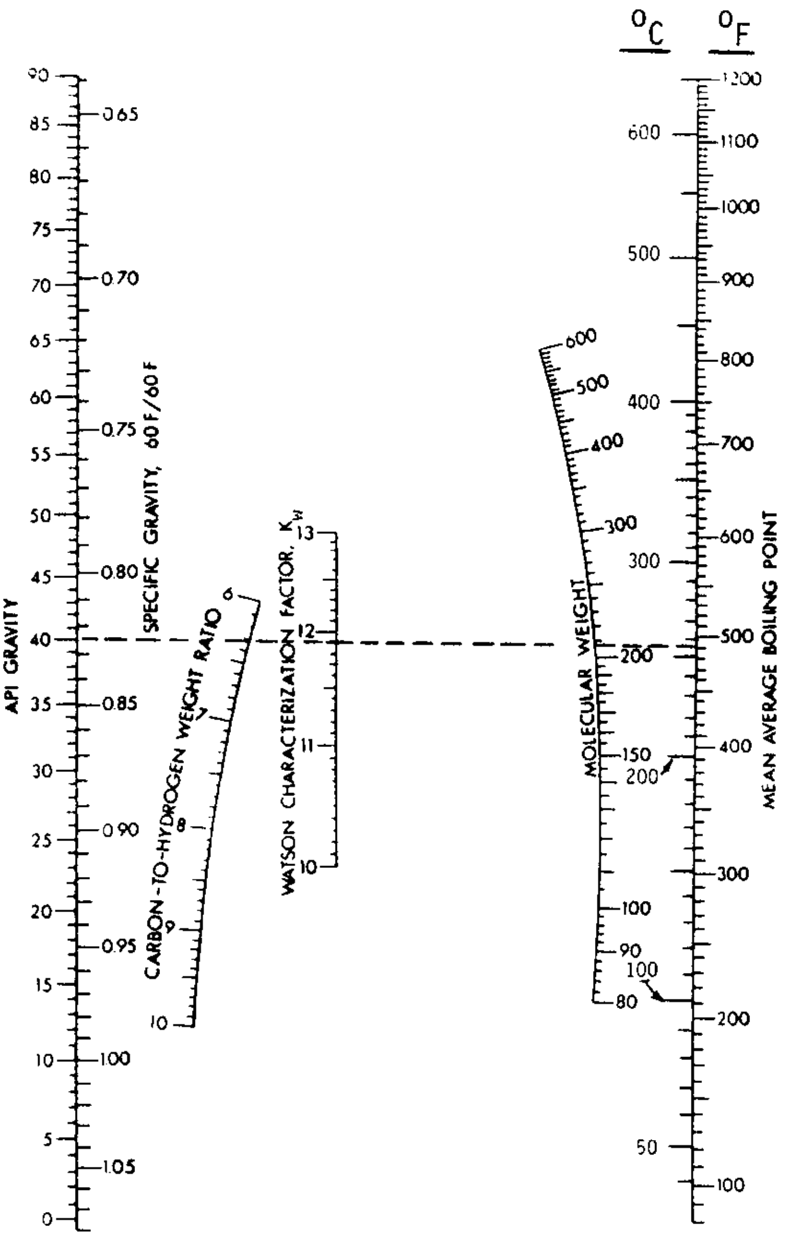


Figure 3.7 Correlations Relating Various Characterization Variables and Liquid Properties

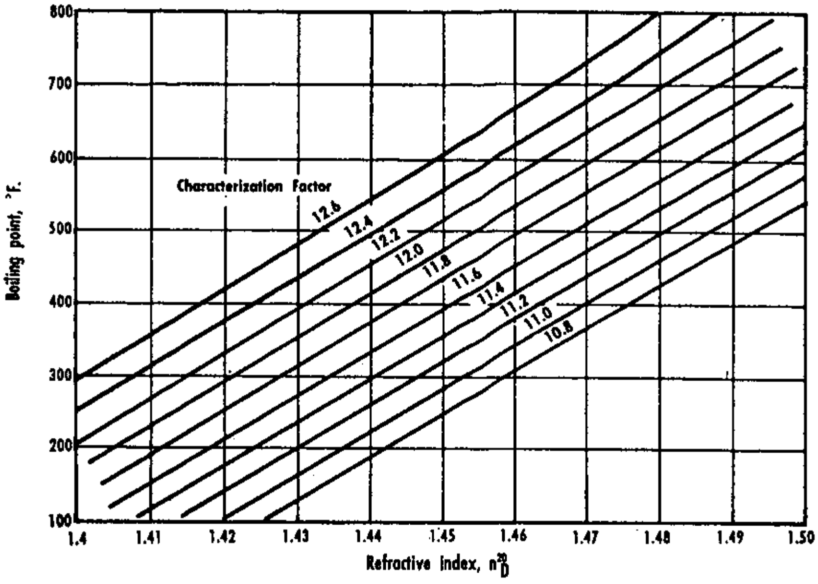


Figure 3.8 **Relationship Between Refractive Index, Characterization Factor, and Boiling Point**

LIQUID DENSITY

Prediction of liquid density is a fundamental calculation. The complexity of this calculation depends on the presence of methane, ethane, CO_2, sulfur compounds and non-paraffins.

Units of Density

Wherever practical we will show liquid density as relative density throughout this book to simplify unit conversions. *Specific gravity*, by definition, was the density of a liquid divided by the density of pure water. In new SI standards, specific gravity is replaced by the term *relative density*.

At 15°C [60°F] the density of water is 1.000 95 g/cm^3; for engineering calculations 1.0 g/cm^3 will suffice at any ambient air temperature. As a practical matter, the error in using 1.0 g/cm^3 is trivial. For this case the density, ρ, for any fluid is

$$(\text{g/cm}^3) = \gamma \qquad (\text{kg/m}^3) = 1000\,\gamma \qquad (\text{lb/ft}^3) = 62.4\,\gamma$$

API Gravity

The density of crude oil is sometimes shown in °API. This term is defined by the equation

$$\gamma = \frac{141.5}{131.5 + {}^\circ\text{API}}$$

$$^\circ\text{API} = \frac{141.5}{\gamma} - 131.5 \tag{3.18}$$

API Barrel

The word "barrel" traditionally refers to the term "API barrel." By definition, 1 API barrel = 42 U.S. gallons. The reference temperature for an API barrel is 15.5°C [60°F].

If you are using a density term where the volume is expressed in barrels (bbl), the following conversion factors are useful:

$$1 \text{ API bbl} = 42 \text{ U.S. gallons} = 35 \text{ U.K. (Imperial) gallons} = 5.61 \text{ ft}^3$$

$$= 0.159 \text{ m}^3 = 159 \text{ liters}$$

General Correlations

A series of general correlations are available that are suitable for most routine applications and can be performed manually.

One simple approach is to use the Law of Ideal Solutions which says that the properties of a mixture are the sum of the properties of its individual parts (components). An equation like Kay's Rule can be used for this purpose. This approach is fine for finding apparent molecular weight but has more limited application when the liquid contains substantial quantities of methane and ethane. The density of liquids containing these components is lower than that predicted by the ideal solution law and by correlations developed for liquids containing primarily propanes and heavier.

The law of ideal solutions, called Amagat's Law, is given by Equation 3.19.

$$\nu_{\text{mix}} = \Sigma\, x_i \nu_i \tag{3.19}$$

Where: ν_{mix} = molar volume of the mixture
x_i = mol fraction of each component
ν_i = molar volume of each component

The accuracy required for the density calculation depends on the use. From an energy viewpoint, a liquid can be regarded as incompressible (Chapter 7) with pressure; pressure has little effect on the liquid energy level. For custody transfer between seller and buyer the effect of P and T on density is important. In most engineering calculations the order of magnitude effect of P and T is important but a high degree of precision is not. The correlations which follow have proven useful for the type of systems discussed herein.

The molecular weight and specific gravity of the heaviest fraction in a mixture must be obtained from the analysis. As noted in a later example, one could use the ideal solution law to find apparent molecular weight and apparent specific volume of the mixture and then divide one by the other to find apparent density.

Relative Density Versus Molecular Weight

There is no specific correlation between these two for liquids. However, an approximate correlation is useful for preliminary correlations on systems where detailed data are not available.

A plot of molecular weight vs. relative density depends on the molecular species involved and the composition of the mixture. Figure 3.9 is an approximate correlation. The lower line is for liquids recovered from natural gas. The upper curve is for naturally occurring reservoir liquids which have a Watson Characterization factor of about twelve. The values shown are consistent with Figure 3.7.

General Nomographs

There is a series of nomographs available which is useful for predicting the density of liquids containing no significant amounts of methane and ethane. They apply primarily for crude oils separated near low pressure, or for natural gas liquids that have been deethanized or separated at ambient temperatures at 3.5 MPa [500 psia] or below. These nomographs will usually yield results that are within 3-5% for the liquids described above, provided the temperature is below 0.9 T_c.

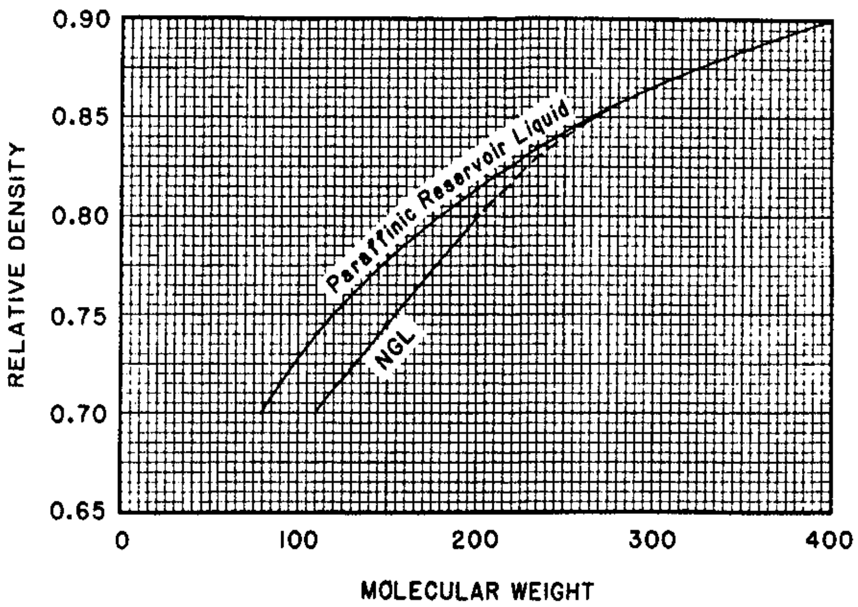


Figure 3.9 **Approximate Correlation between Molecular Weight and Relative Density for Hydrocarbon Liquids**

Figure 3.10 is a modified version of an API correlation. The middle section accounts for composition (K_w) and standard density (or relative density). The intersection of these two lines fixes a point on the grid. A straight line is drawn through this point from the applicable temperature to find density, at that temperature.

In the example shown, the density is found as 0.67 g/cm^3 [41.8 lb/ft^3] at 41°C [106°F] for a liquid with a standard relative density of 0.7 and $K_w = 12.0$.

The line labelled iso- and n-paraffins may be used to find the density of pure components at any temperature.

Figure 3.11 is a similar correlation from Reference 3.4, modified for use with metric units. A point on the middle grid is located easily from the intersection of any two lines. In most cases K_w and API gravity are used. The example is for a liquid with a K_w of 11.0 and an API Gravity of 40, at 6.9 MPa [1000 psia] pressure. The result is 0.61. At atmospheric pressure the result would be 0.56. Thus, the pressure has increased relative density. Figure 3.11 possesses the same limitations as Figure 3.10.

Figure 3.12 is a simple correlation for estimating the density of a saturated paraffin hydrocarbon liquid. Since a liquid leaving a gas-liquid separation is saturated (at its bubblepoint), at the temperature and pressure of the separation, this correlation is useful. It should be used with caution if the liquid contains significant amounts of aromatics and/or naphthenes.

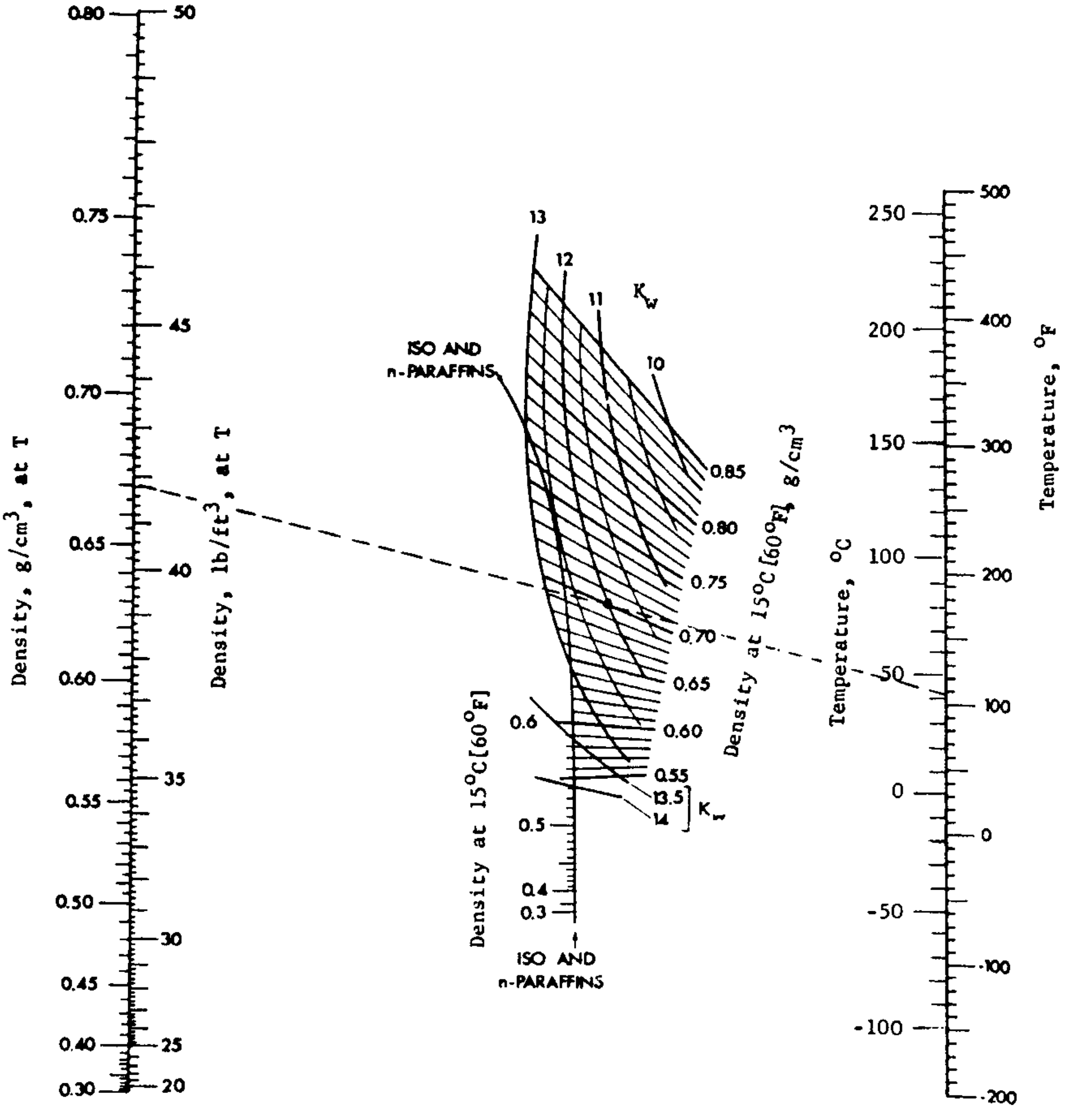


Figure 3.10 Density - Temperature Correlation for Liquids

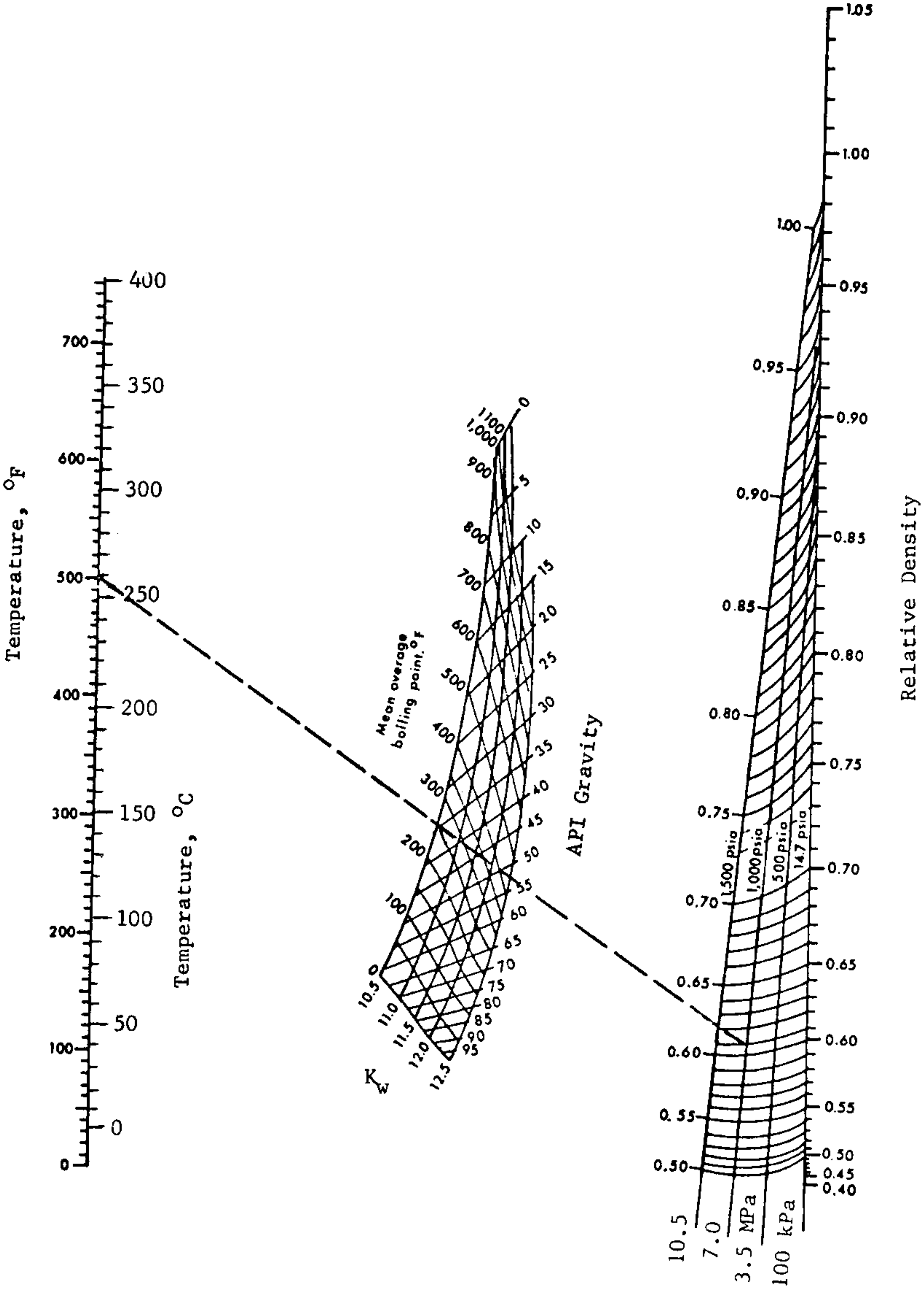


Figure 3.11 Relative Density of Hydrocarbon Mixtures

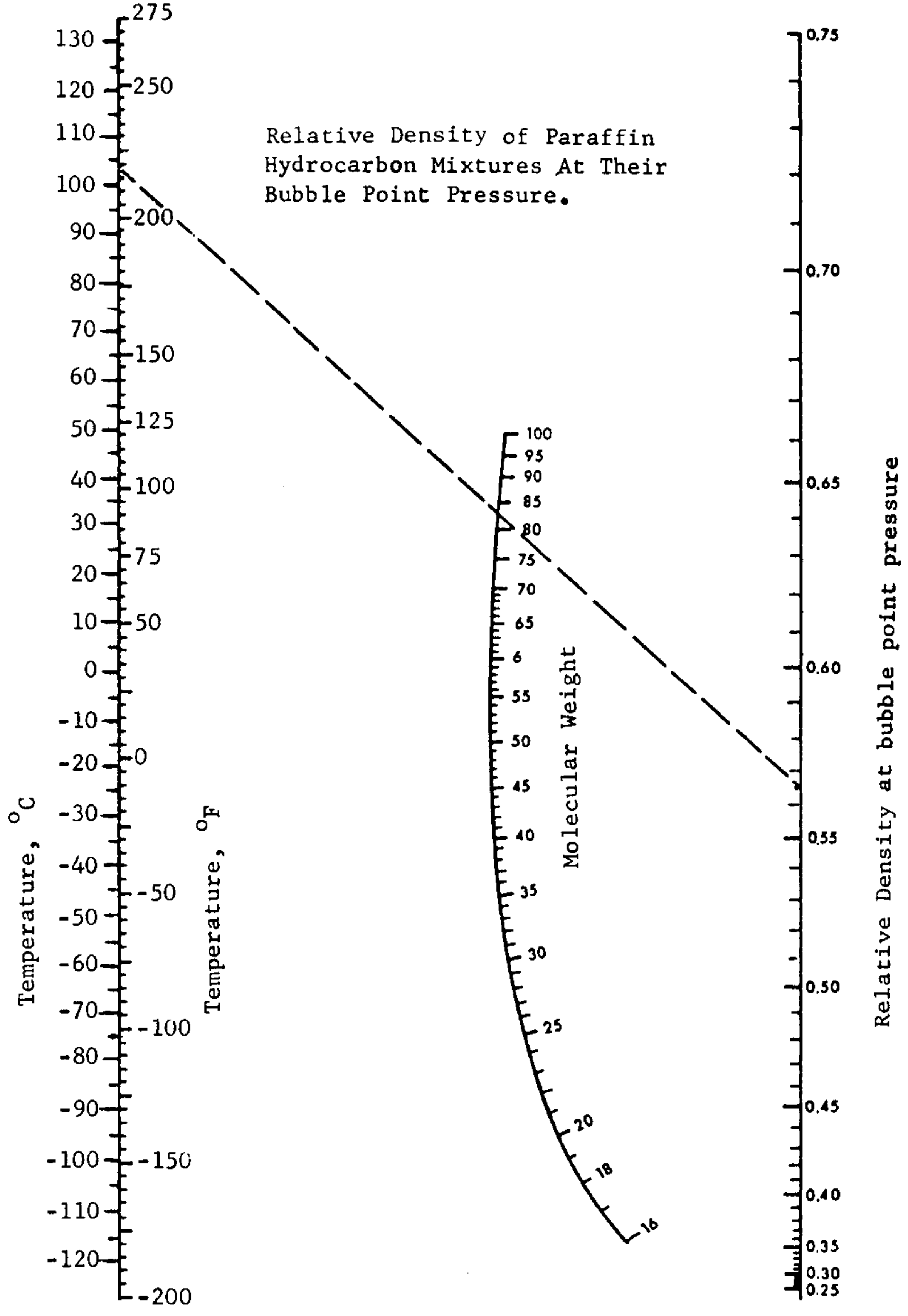


Figure 3.12 Relative Density of Hydrocarbon Mixtures at their Bubble Point Pressure

Corresponding States Methods

Several corresponding states models have been proposed for prediction of liquid densities. Three methods widely used in the industry are:

1. API Method(3.14)
2. Modified Rackett Equation(3.15)
3. Hankinson-Brobst-Thomson [COSTALD](3.16, 3.17)

Of these, only method one readily lends itself to hand calculation. Figure 3.13 is a plot of C, a density factor as a function of P_r and T_r. The density factor is used to correct the known density, ρ_1, at one condition to a density, ρ_2, at another condition using the equation

$$\rho_2 = \rho_1 \left(\frac{C_2}{C_1} \right) \tag{3.20}$$

In Figure 3.13 and Equation 3.20 the critical (or pseudocritical) properties are used to find the reduced T and P at conditions (1) and (2). Values of C_1 and C_2 are then found from Figure 3.13. The quantity ρ_1 could be found from the Law of Ideal Solutions or one of the nomographs discussed above. The critical-values could be found from Tables 3.1 or 3.2 for a pure substance or from Kay's rule for a mixture.

The API method can be used to predict densities of hydrocarbon liquid mixtures when the mixture is composed of molecules of similar size and shape, e.g. C_3-C_4, C_5-C_6, etc. It is less accurate for mixtures containing a wider range of molecules, e.g. C_1-C_7.

The modified Rackett equation can be used to estimate the density of saturated liquid. For pure components the basic equation is

$$v_s = \frac{RT_c}{P_c} Z_{RA}^{[1 + (1 - T_r)^{2/7}]} \tag{3.21}$$

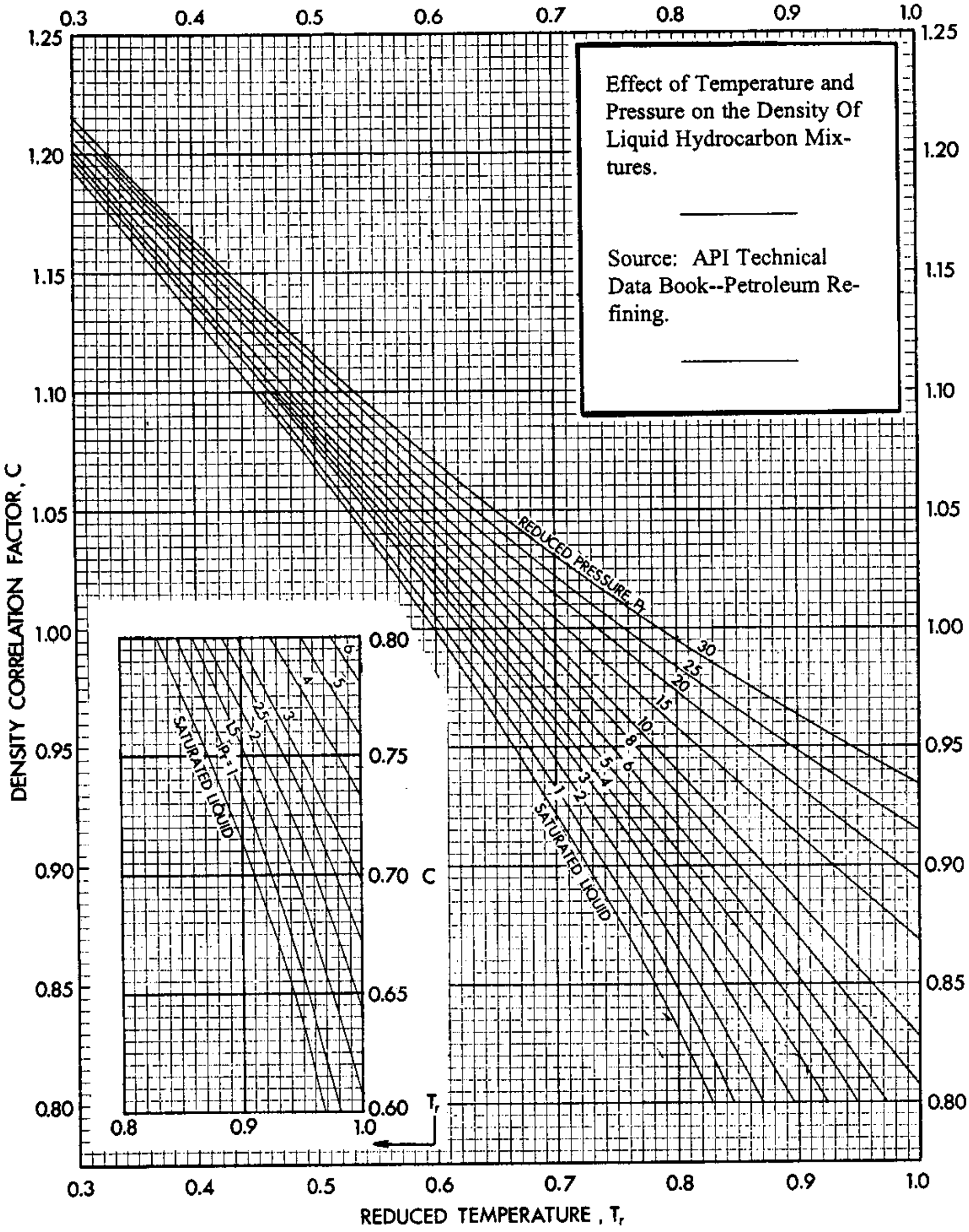
			Metric	English
Where:	v_s =	molar volume of saturated liquid at T	m^3/kmol	ft^3/lb-mol
	R =		8.314	10.73
	T_c =	critical temperature	K	°R
	P_c =	critical pressure	kPa	psia
	Z_{RA} =	empirical constant unique to each component (if Z_{RA} is not known it can be approx. by Z_c)		
	T_r =	reduced temperature		

Equation 3.21 can also be applied to mixtures, however, the mixing rules for the determination of T_c are quite complex and require computer solution.

The Hankinson-Brobst-Thomson method is sometimes referred to as the COSTALD method (Corresponding States Liquid Densities). For pure components it employs the basic equation

$$\frac{v_s}{v^*} = v_r^o \left[1 - \omega_{SRK}\, v_r^\delta \right] \tag{3.22}$$

Where: v_s = molar volume of saturated liquid at T
v^* = an empirical constant similar to Z_{RA} in Equation 3.20
It is typically within 1 to 4% of v_c
ω_{SRK} = SRK acentric factor
v_r^o and v_r^δ = reduced volume factors which depend on T_r



Effect of Temperature and Pressure on the Density Of Liquid Hydrocarbon Mixtures.

Source: API Technical Data Book--Petroleum Refining.

Figure 3.13 **Effect of Temperature and Pressure on the Density of Hydrocarbon Liquids**[3.11]

Values for υ^*, ω_{SRK}, and equations for υ_r^o and υ_r^δ may be found in Reference 3.16. Equation 3.22 is also applicable to mixtures. Special mixing rules are used to determine υ^* and T_c for the mixture. These are too complex to permit hand calculation. Binary interaction parameters have been used to improve the accuracy on specific mixtures.(3.18, 3.19)

The COSTALD method can also be used to predict densities of compressed liquids.(3.17)

Example 3.4: Estimate the density of an EP_{mix} (80% C_2, 20% C_3) at 10 000 kPa [1450 psia] and 10°C [50°F] using Figure 3.13 and Equation 3.19.

Step 1. Determine the density at reference (saturated) conditions using Eqn 3.19. Pure component densities may be found from Table 3.2.

From Table 3.2(a)

$$\gamma_{C_2} = 0.3581 \quad ; \quad \rho = 358.1 \text{ kg/m}^3$$

$$\gamma_{C_3} = 0.5083 \quad ; \quad \rho = 508.3 \text{ kg/m}^3$$

From Equation 3.18

$$v_{mix} = \sum v_i x_i = \sum \frac{MW_i\, x_i}{\rho_i}$$

$$v_{mix} = \frac{30.07(0.8)}{358.1} + \frac{(44.097)(0.2)}{508.3} = 0.084\ 53 \text{ m}^3/\text{kmol}$$

$$MW_{mix} = (30.07)(0.8) + (44.097)(0.2) = 32.875$$

$$\rho_{mix} = \frac{MW_{mix}}{v_{mix}} = \frac{32.875}{0.084\ 53} = 388.9 \text{ kg/m}^3$$

Step 2. Find C_1 – Apply Kay's rule and Figure 3.1 to determine P_c' and T_c'

$$P_c' = (0.8)(4.88) + (0.2)(4.25) = 4.75 \text{ MPa}$$

$$T_c' = (0.8)(305) + (0.2)(370) = 320 \text{ K}$$

Reference conditions at 15°C and saturation

$$P_{r_1} = \text{saturated liquid}$$

$$T_{r_1} = \frac{288}{320} = 0.90$$

From Figure 3.13, $C_1 = 0.715$

Step 3. Find C_2

$$P_{r_2} = \frac{10.0}{4.75} = 2.1$$

$$T_{r_2} = \frac{283}{320} = 0.884$$

From Figure 3.13, $C_2 = 0.79$

$$\rho_2 = \rho_1\left(\frac{C_2}{C_1}\right) = 388.9\left(\frac{0.79}{0.715}\right) = 430 \text{ kg/m}^3$$

Effect of Methane and Ethane

A liquid containing significant quantities of methane and ethane will have a lower density due to the high kinetic energy of the molecules. Figures 3.15-3.14 offer a relatively easy, reasonable correlation for such mixtures.(3.20)

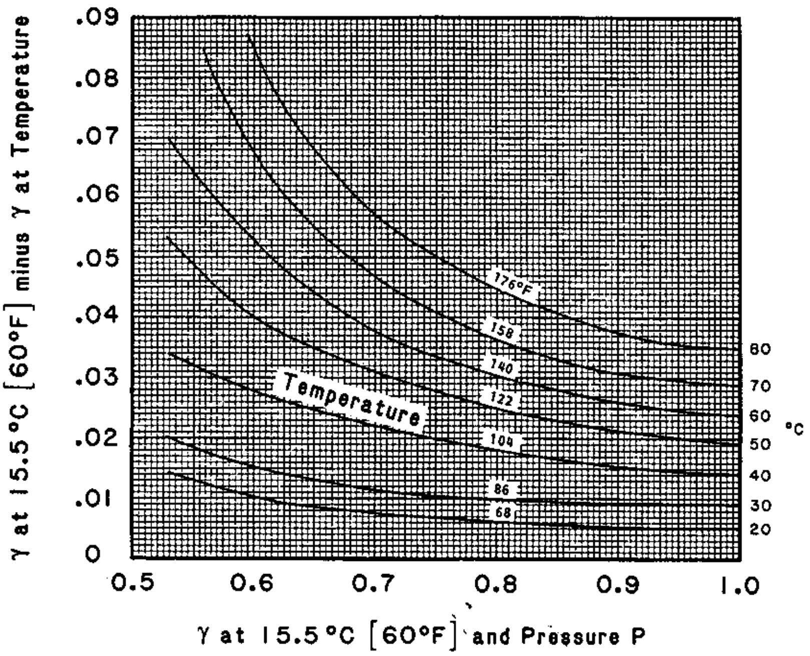


Figure 3.14 Temperature Correction for Relative Density in Figure 3.16

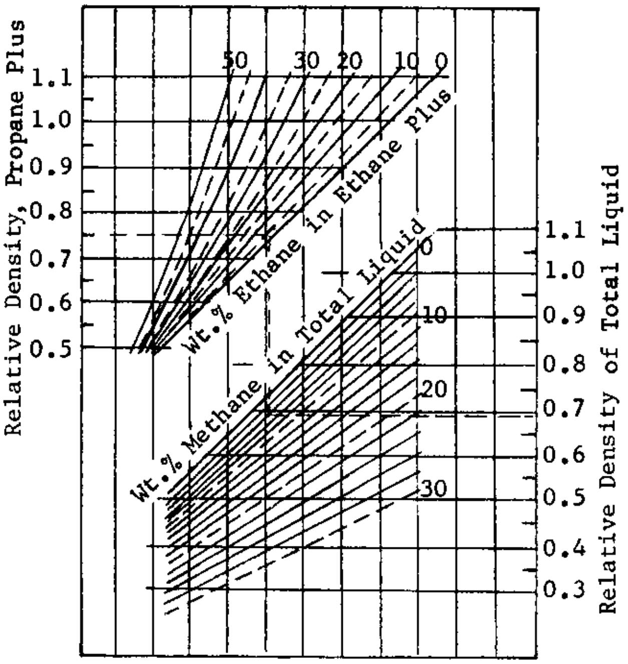


Figure 3.15 Relative Density of Liquids Containing Methane and Ethane at Atmospheric Pressure and 15°F

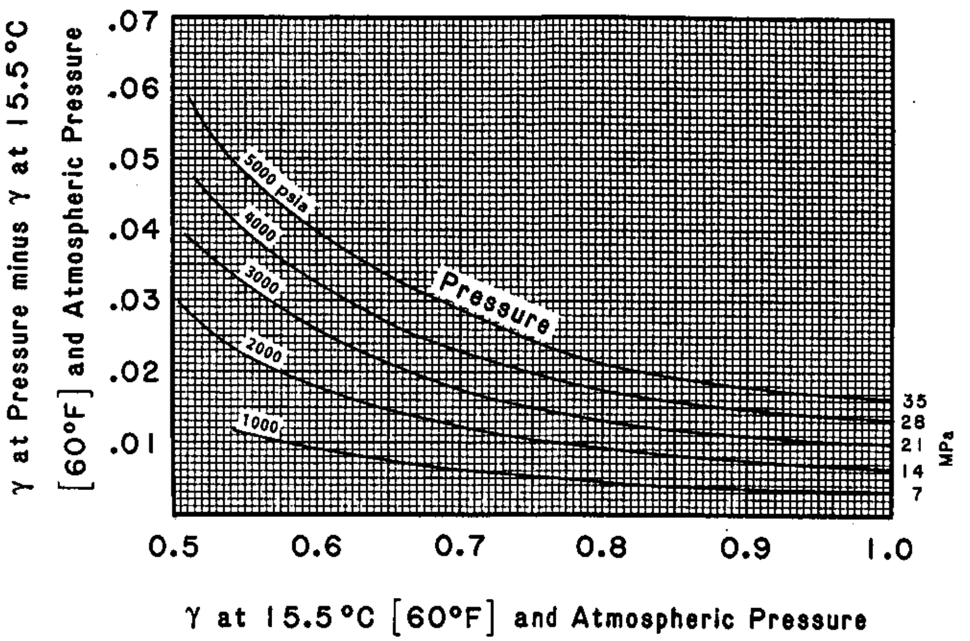


Figure 3.16 **Pressure Correction for Relative Density in Figure 3.15**

From a known liquid analysis the first step is to calculate the *weight percent* ethane in the ethane plus fraction and *weight percent* methane in the total liquid.

This correlation is the first of several in this book that involves conversion from one type of percentage to another. The relative molecular weight is the key to the conversion between mol and weight fractions (or percentages). In Example 3.5, Column (2) shows the number of mols of each component per mol of mixture. Multiplying each of these quantities by its relative molecular weight (Column 3) gives the corresponding mass (Column 4).

The liquid in this example is typical of that from low temperature processing. Over 30% is methane and ethane. A substantial part of the liquid must be vaporized, the amount depending on whether ethane is recovered as a liquid or not.

The relative density of 0.69 is at 15°C and atmospheric pressure. What would be the density at 9 MPa [1305 psia] and 77°C [170°F]? The result from Figure 3.15 is entered on the abscissa of Figure 3.16. Proceed vertically to the pressure and then horizontally to the left ordinate. The ordinate value for our example (0.0085) is *added* to the atmospheric value (0.69) to obtain the value corrected for pressure (0.69 + 0.0085 = 0.6985). This pressure corrected value then is entered on the abscissa of Figure 3.14. Proceed vertically to the temperature and horizontally to the left ordinate. This ordinate value for our example (0.054) is *subtracted* to obtain the temperature correction (0.6985 – 0.054 = 0.6445).

The relative molecular weight of our example liquid is 86.68. From Figure 3.9 the relative density (by visual extrapolation) would be about 0.67. From Figure 3.12 it would be about 0.61. As a general rule, a liquid containing large amounts of methane and ethane will have a lower density than most oil type correlations predict. If you need to estimate the density of a liquid probably containing significant amounts of methane and ethane but do not have an analysis, Figure 3.12 should give the best estimate among those shown.

Example 3.5: Calculate the density of the liquid shown in Column 1 and 2 below.

(1) Component	(2) x_i	(3) Mol Wt	(4) = (2)(3) Wt of 1 mol	(5) γ	(6) = (4)/(5) Vol of Comp.
C_1	0.237	16	3.79		
C_2	0.124	30	3.72		
C_3	0.128	44	5.63	0.508	11.08
iC_4	0.020	58	1.16	0.563	2.06
nC_4	0.075	58	4.35	0.584	7.45
iC_5	0.021	72	1.51	0.624	2.42
nC_5	0.023	72	1.66	0.631	2.63
C_6	0.056	86	4.82	0.664	7.26
C_{7+}	0.316	190*	60.04	0.823	72.95
	1.000		86.68		105.85

* C_{7+} = 190 MW and $\gamma = 0.823$ from analysis

$$\gamma_{C_3+} = \frac{86.68 - 3.79 - 3.72}{105.85} = 0.748 \quad , \quad \text{Wt\% ethane in } C_2+ = \frac{3.72}{86.68 - 3.79}(100) = 4.49\%$$

$$\text{Wt\% } C_1 = \frac{3.79}{86.68}(100) = 4.37\%$$

From Figure 3.15, $\gamma = 0.69$

Density = 0.69 g/cm^3 , 690 kg/m^3 , or (0.69)(62.4) = 43 lb/ft^3

Summary of Liquid Density Correlations

In general, liquid density correlations can be placed in three categories

1. Empirical P and T correlations
2. Corresponding States Methods
3. Equations of State

Empirical P and T correlations are used to correct a known density to actual process conditions. They are frequently used on crude oils and condensates for which no analysis is available, but for which a stock-tank density or API gravity may be known. Figures 3.10 and 3.11 fall into this category.

Corresponding States Methods are also empirical and designed to correct a known density to actual process conditions using the corresponding states principle. In the API method the "known" density is either measured or calculated from Amagat's Law. In both the Rackett and COSTALD methods the "known" density is essentially the critical density. These methods require a liquid composition.

Equation of State methods may also be used to estimate liquid densities. In general, the cubic equations of state (vdW, SRK, PR) are less accurate than the corresponding states models. The BWRS equation of state is considered more accurate for liquid density predictions.

For liquids of know composition, the COSTALD method is probably the most general and the most accurate of the methods presented. For crude oils and condensates Figures 3.10 and 3.11 provide acceptable results for engineering calculations.

VISCOSITY

Viscosity is another important property of a hydrocarbon fluid used primarily in flow calculations. It is that number which represents the drag forces caused by the attractive forces in adjacent fluid layers. It might be considered the internal friction between molecules, separate from that between the fluid and the solid wall.

Since the intermolecular forces are a function of the distance between molecules, viscosity depends on whether the fluid is a liquid or gas at system temperature and pressure. The viscosity of a liquid is much higher than that for a gas at the same conditions of pressure and temperature. As with density, it is convenient to consider liquid and gas viscosities separately.

Units of Viscosity

The basic petroleum viscosity unit has been the poise, or more commonly the centipoise. The poise is a measure of absolute (dynamic) viscosity. It is a derived metric unit: 1 poise = 1 (dyne)(sec) per cm^2.

$$1 \text{ centipoise (cp)} = 0.01 \text{ dyne s/cm}^2 = 0.001 \text{ Pa·s}$$

$$= 0.000\ 672 \text{ lbm/ft–sec}$$

The kinematic viscosity is the absolute viscosity divided by density. The usual unit is the stoke or centistoke.

$$1 \text{ centistoke} = 0.01 \text{ cm}^2/\text{s} = 1.0 \times 10^{-6} \text{ m}^2/\text{sec}$$

The correlations which follow have proven useful and reliable for manual calculations involving the viscosity term. Chapter 10 will show the application of viscosity to fluid flow.

GAS VISCOSITY

The viscosity of natural gas at 100 kPa [14.5 psia] may be estimated from Figure 3.17.(3.21) Corrections are provided for N_2, CO_2 and H_2S (where G = gas relative density). The molecular weight is the apparent molecular weight found using Kay's Rule or from the relative density at standard conditions.

One may estimate gas viscosity for a mixture from the following equation:

$$\mu_m = \frac{\Sigma \mu_i y_i \sqrt{M_i}}{\Sigma y_i \sqrt{M_i}} \tag{3.23}$$

Where: μ_m = mixture viscosity
μ_i = viscosity of each component
y_i = mol fraction of each component
M_i = molecular weight of each component

This equation usually agrees with Figure 3.17 within acceptable limits.

The correction for temperature and pressure is made from Figure 3.18, where T_r and P_r are found from the critical values obtained by Kay's Rule. In this figure the value on the left hand ordinate (μ/μ_1) is the ratio of the viscosity at actual conditions (μ) divided by the viscosity found from Figure 3.17 (μ_1). Notice that gas viscosity is very low, which is an indication of the relatively large distances between molecules. References 3.22-3.25 present other high pressure gas viscosity correlations.

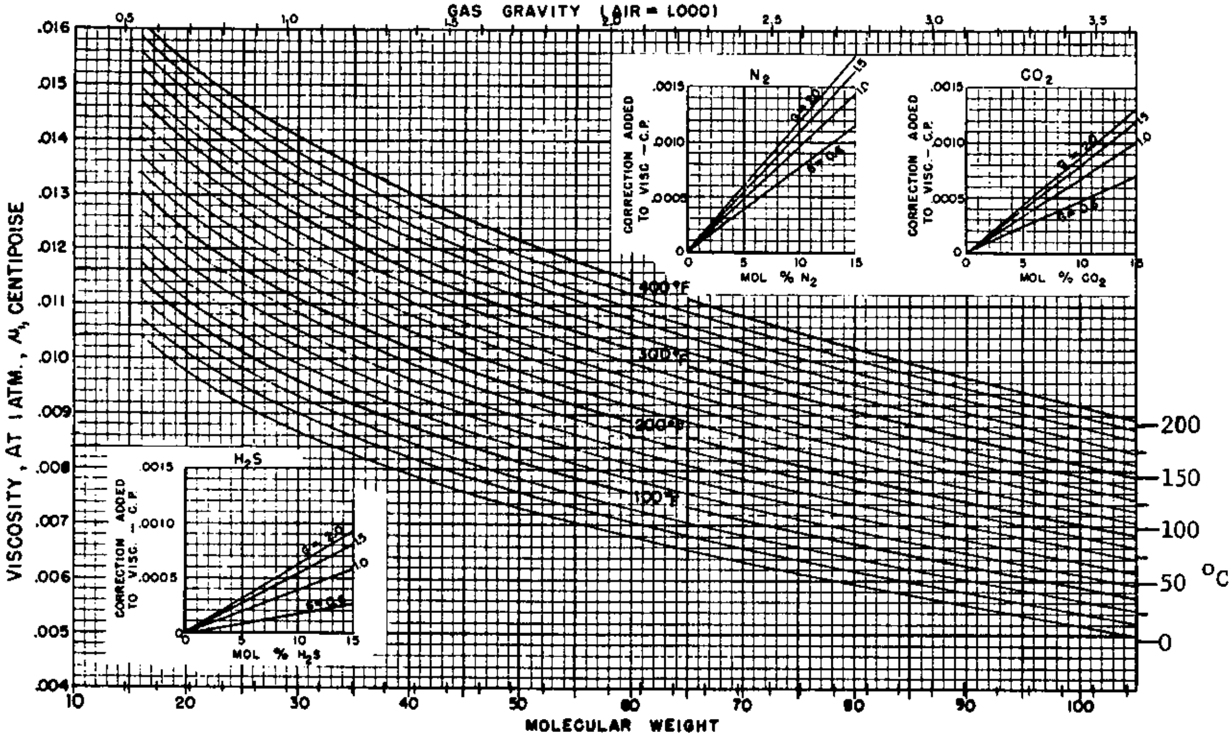
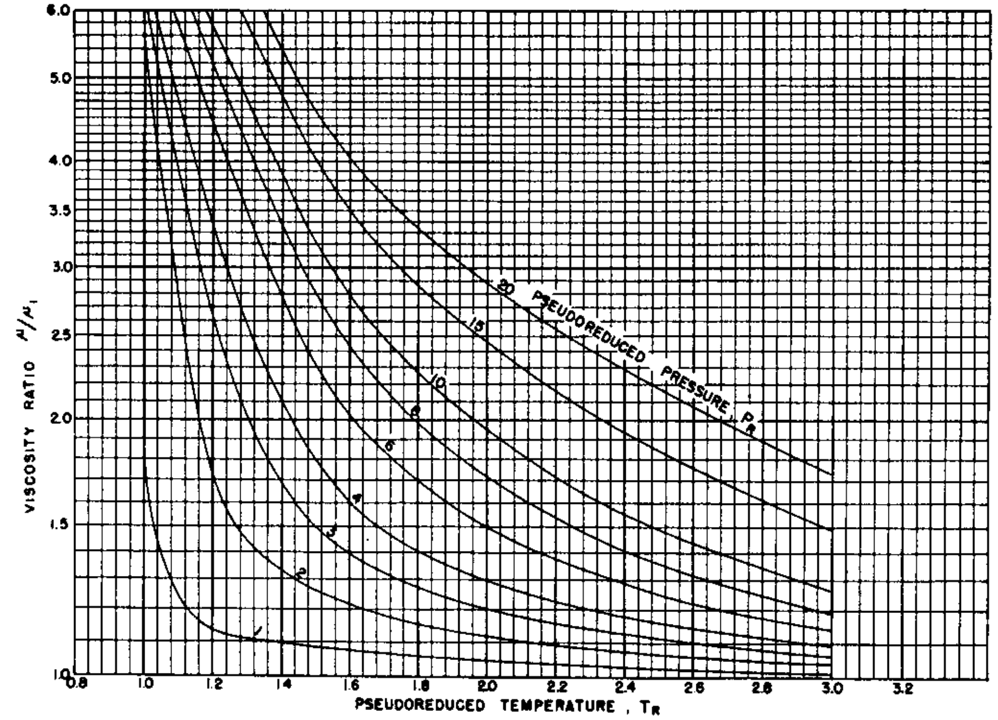


Figure 3.17 Viscosity of Paraffin Hydrocarbon Gases at Atmospheric Pressure



μ_1 = Visc @ 1 atm.

Figure 3.18 Pressure Correction for Viscosity from Figure 3.17

LIQUID VISCOSITY

The prediction of liquid viscosity is somewhat analogous to the prediction of density. One can use P-V-T equations of state and combination rules or general correlations based on physical parameters like molecular weight, EMR, etc. Many of the former are not suitable for manual calculations. References 3.26-3.32 present such methods and those not included in the discussion herein.

Figure 3.19 shows the viscosity at atmospheric pressure for common light paraffin hydrocarbons, as a function of temperature. One can use such data to estimate the viscosity of mixtures using the equation

$$\mu_m = \left[\Sigma x_i (\mu_i)^{1/3}\right]^3 \tag{3.24}$$

Where: μ = viscosity of mixture in centipoise, μ_i = component viscosity
x_i = mol fraction of each component in the mixture

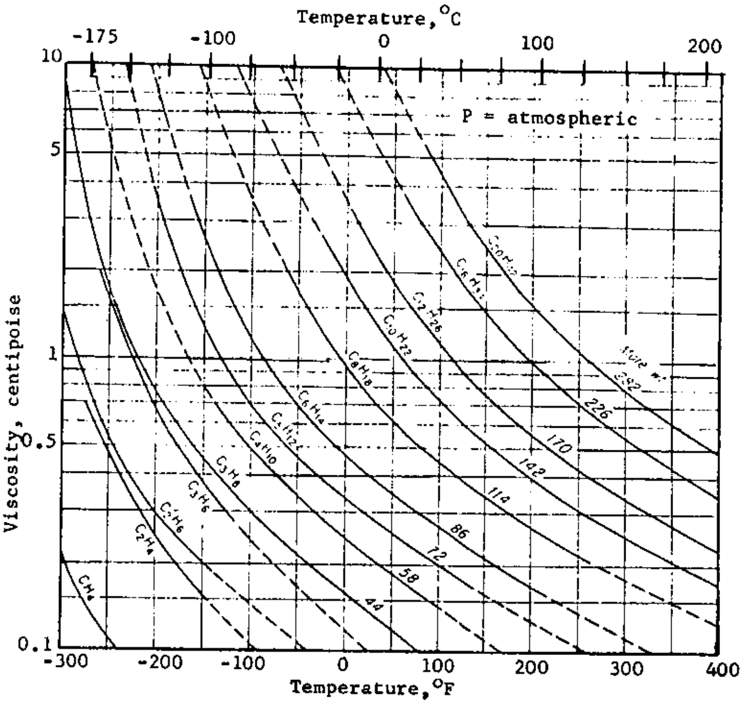


Figure 3.19 Viscosity of Pure Paraffin Hydrocarbons

The viscosity of crude oils possessing an API gravity greater than 30°API ($\gamma = 0.88$) can be estimated by the equation

$$\log \mu = a - (0.035)(°\text{API}) \tag{3.25}$$

Where: a = values at right
°API = $(141.5/\gamma) - 131.5$
μ = viscosity, cp

°C	a
38	2.05
54	1.83
71	1.55
88	1.30
104	1.08

A more detailed correlation of data for crude oils is found in Reference 3.30.

Correlations for prediction of liquid viscosity possess a general reliability of 13-15%. This range offers no particular practical problem in the type of calculations where viscosity is used.

SURFACE TENSION

A molecule on the surface of a liquid is subject to an inward force as a result of the attraction between molecules. This surface molecule tends to adjust itself to a minimum surface area causing the droplets to assume a spherical shape. Work is necessary to extend this surface or to bring molecules from the bulk of the liquid into the surface. This work has been expressed traditionally in ergs/cm^2 and is a measure of what is called *free surface energy.* If one thinks of the attractive forces as a film over the liquid surface one can speak of the result as a surface tension, traditionally expressed as dynes/cm.

The dyne is not a basic unit of force in the SI system, the newton (N) replacing it.

$$1 \text{ dyne} = 10^{-5} \text{ newtons}$$

$$1 \text{ dyne/cm} = 10^{-3} \text{ N/m}$$

One simple approach for calculating surface tension is found using Equation 3.26.

$$\sigma = C(M)^n \tag{3.26}$$

Where:
σ = surface tension, dynes/cm
M = molecular weight
C = use table
n = use table

Compound	C	n
Paraffin hydr.	2.26	0.47
Aromatic hydr.	23.0	0.05
Alcohols	14.5	0.12
Ketones	12.0	0.17

The values of σ from this equation apply at 20°C and atmospheric pressure. One can correct for temperature by the equation

$$\sigma'/\sigma = \left[\frac{(1 - T_r')}{(1 - T_r)}\right]^{1.2} \tag{3.27}$$

Where: σ' and T_r' are at conditions other than 20°C

Another approach$^{(3.33)}$ uses the equation

$$\sigma^{0.25} = (P/M)(\rho_L - \rho_v) \tag{3.28}$$

Where:
σ = surface tension, dynes/cm (1 dyne/cm = 0.001 N/m)
P = parachor
M = molecular weight of the liquid
ρ_L = liquid density, g/cm3
ρ_v = vapor density, g/cm^3

The quantity P may be estimated as follows:

P = 40 + (2.6)(MW) for paraffins, and mixtures to a mol wt of 100
P = 300 + (2.25)(MW – 100) for mixtures having a mol wt above 100

Equation 3.28 is only valid below a reduced temperature of 0.85.

Surface tension is affected by temperature and the composition of the liquid and vapor around a drop. Pressure has an indirect effect only due to its effect on composition and density. Reference 3.33 also presents a method for estimating the surface tension of mixtures.

GENERAL PREDICTION OF FLUID PROPERTIES

The use of equations of state and/or corresponding states methods for prediction of fluid behavior has increased dramatically with the advent of computers. Physical, phase behavioral and thermodynamic properties are produced within the same modeling format. The resultant equations are replacing graphical correlations.

References 3.27 and 3.34-3.36 are early work using a series of coefficients to find the density of mixtures. Sarem and Campbell*(3.37)* proposed a similar approach for gas condensates near the critical region.

As mentioned earlier, the COSTALD equation has been used to successfully predict density for a large number of hydrocarbons. This same approach has been utilized to estimate the density of spiked crudes.*(3.38)*

Pan and Maddox *(3.39)* present a series of useful equations for calculating both physical and thermodynamic properties of saturated liquids and vapors.

A very thorough presentation has been made for liquids containing only heptanes and heavier molecules based on data available from a TBP distillation of this liquid.*(3.40)* A number of properties are addressed but the following equations are shown for molecular weight and density.

$$M = (1.66 \times 10^{-4})(T_b)^{2.20}(\gamma)^{-1.02} \qquad \rho = (9.84 \times 10^{2})(T_b)^{0.002}(\gamma)^{1.006} \tag{3.29}$$

Where:
M = molecular weight, kg/kmol
T_b = avg. boiling point, K
γ = relative density at 15.5°C [60°F]
ρ = liquid density, kg/m^3

Using correlations like these in equation form for computer solutions is most convenient. But, it should be remembered that they also are empirical and limited by the extent of the data used to obtain the empirical constants. The greater the variety of molecular types in the mixture the less reliable the empirical correlation becomes. The interaction of dissimilar molecules cannot totally be represented by a simple empirical equation.

In most cases, though, the error in physical property prediction will have far less effect on system design than specification of flow rates and conditions, errors in analysis, etc., if said prediction is done thoughtfully.

USEFUL CONVERSION FACTORS FOR CHAPTER 3

°F = (9/5) °C + 32 , °C = (°F − 32)(5/9) , °R = 1.8 K , K = 0.555°R

100 000 N/m^2 = 100 kPa = 14.50 psia = 1.02 kg/cm^2 = 0.987 atm

1 lb_m = 0.454 kg , 1 kg = 2.205 lb_m

1 ft^3 = 28.3 L = 0.028 m^3 ; 1 m^3 = 35.31 ft^3 = 1000 L

1 API bbl = 159 L = 5.61 ft^3 = 0.159 m^3

γ (liquid) = Density of liquid/Density of water...at some P and T. = 141.5/(131.5 °API)

Density of water ≈ 1.0 g/cm^3 ≈ 1000 kg/m^3 ≈ 62.4 lb/ft^3

γ (gas) = (Mol Wt gas)/(Mol Wt air) = MW gas/28.97
= Density of gas/Density of air … at standard P and T

Density of air at 15°C and 100 kPa = 1.21 kg/m^3

Density of air at 60°F and 1 atm (14.7 psia) = 0.0764 lb/ft^3

1 cp = 0.01 dyne·s/cm^2 = 0.001 N·s/m^2 = 0.001 kg(m·s) = 0.000 672 lbm/(ft−s)

1 centistoke = cp/ρ = 0.01 cm^2/s = 1.0 (E−06)m^2/s

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4

QUALITATIVE PHASE BEHAVIOR

In this book we are concerned primarily about the application of energy to achieve desired goals. The energy possessed by any substance depends on its phase. There are three different familiar phases ... solid, liquid and gas (vapor). You have learned to identify them by sight.

A solid possesses a definite shape and is hard to the touch. It is composed of molecules with very low energy that stay in one place even though they vibrate. There is space between these molecules so a solid is not impenetrable.

A liquid has a definite volume but no definite shape. It will assume the shape of the container in which it is placed but will not necessarily fill that container. The molecules of which the liquid is composed possess more energy than in a solid; enough energy to move from place to place. By virtue of this energy there is more space between molecules. So, a substance is less dense in the liquid form than in the solid form.

A vapor has no definite volume or shape and will fill a container in which it is placed. The molecules have more energy than in the liquid form. In fact, they are very active. Vapor density usually is low enough that one can see through a vapor. As might be expected, the density of a given substance is lower in the vapor phase than in the liquid phase.

Our primary concern is the difference in energy level between phases. If we wish to melt a solid to form liquid we must add energy. If enough additional energy is added, the liquid can be vaporized.

We must know the phase or phases that exist at given conditions of pressure, volume and temperature in order to ascertain the corresponding energy level. To do this, we separate substances into two classifications – *pure substances* (single component systems) and *mixtures* of substances (multi-component systems).

SINGLE COMPONENT SYSTEMS

The word component refers to the number of molecular or atomic species present in the substance. A single component system is composed entirely of one kind of atoms or molecules. We often use the word "pure" to describe a single component system.

Figure 4.1 is a typical phase diagram for a pure substance. It has three axes – P, V and T. It is composed of a series of plane surfaces, each of which represents a given phase or a mixture of phases. We are particularly interested in the two-phase planes:

> BDHG – solid plus liquid, FGIJ – solid plus vapor, and an irregular-shaped plane HCI for the liquid plus vapor. All of these planes are *perpendicular* to the temperature axis.

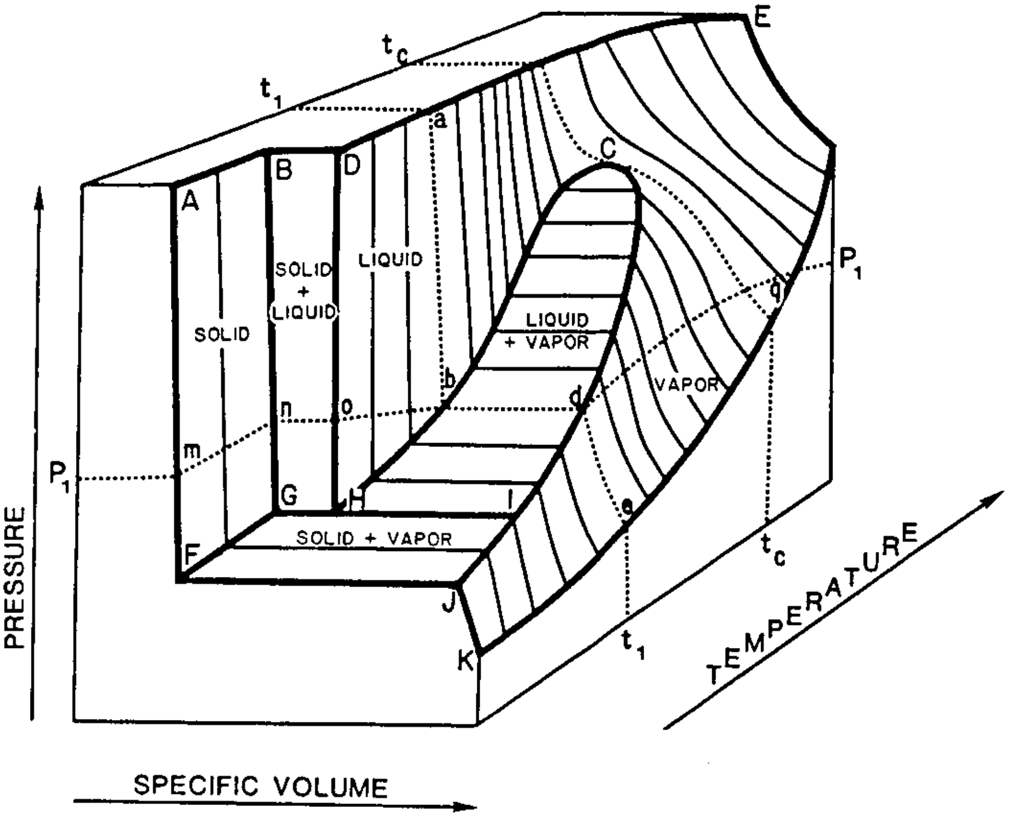


Figure 4.1 **P-V-T Diagram for a Single Component System**

The liquid-only plane is the "cliff" to the left of plane HCI and adjacent to plane BDHG. The vapor-only plane is the "slope" to the right of plane HCI.

Although all planes are of interest, we are primarily concerned with plane HCI, the vapor-liquid region of the phase diagram.

A three dimensional phase diagram like Figure 4.1 is awkward to use. So, we normally draw a projection of this diagram. Both P-T and P-V projections are important.

P-T Plot for a Pure Substance

Since all of the two-phase planes in Figure 4.1 are perpendicular to the T axis, they appear as single lines in a P-T projection like Figure 4.2.

Lines HD, HC, and FH are the equilibrium lines – combinations of pressure and temperature at which the adjoining phases are in equilibrium. At equilibrium one can change phase, at constant pressure and temperature, by simply adding or removing energy from the system. Point H, the triple point, is the only combination of pressure and temperature at which all three phases can exist together.

Along line FH no liquid phase is ever present and solid *sublimes* to vapor. The use of "dry ice" for cooling is an example of this. Line HD is the equilibrium line between solid and liquid. Ice water at 0°C [32°F] and atmospheric pressure occurs on this line. Line HD can have a positive or negative slope depending on whether the liquid expands or contracts on freezing. The energy change occurring along line HD is called the *heat of fusion*. At any P and T along this line the system can be all solid, all liquid or a mixture of the two depending on the energy level. This line could be called the solid-liquid saturation or solid-liquid equilibrium line.

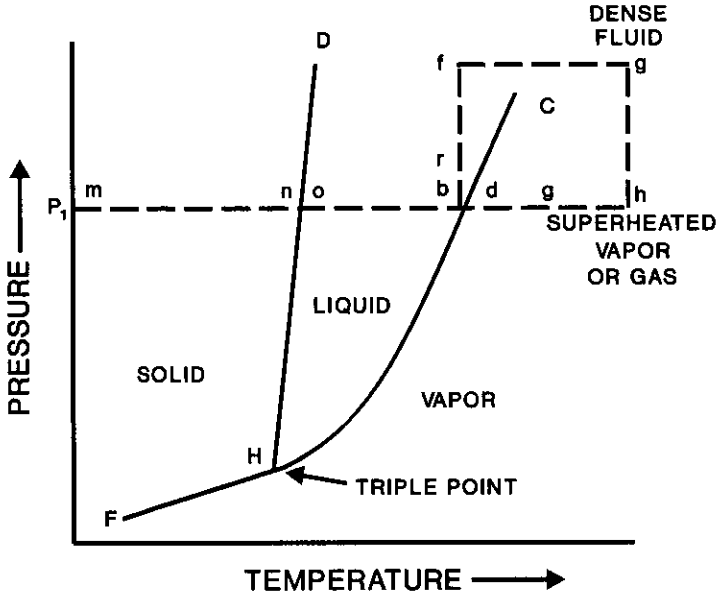


Figure 4.2 **P-T Diagram for a Single Component System**

Line HC is the saturation or equilibrium curve between vapor and liquid. It starts at the triple point and terminates at the critical point "C." The pressure and temperature conditions at this latter point are known as critical temperature (T_c) and critical pressure (P_c). At this point the properties of the liquid and vapor phases become identical. For a pure substance the critical point can be defined as that point above which liquid cannot exist as a unique separate phase. Above P_c and T_c the system is oftentimes referred to as a dense fluid to distinguish it from normal vapor and liquid.

Line HC is often referred to as the *vapor pressure curve.* Such vapor pressure curves are available from many sources. Line HC is also the bubblepoint and dewpoint curve for the pure substance (See Chapter 5).

In Figure 4.2, consider a process starting at pressure P_1, and proceeding at constant pressure. From "m" to "n" the system is entirely solid. The system is all liquid for the segment o-b. At "b" the system is a *saturated liquid* – any further addition of energy will cause vaporization at constant pressure and temperature. At "d," the system is in the *saturated vapor* state. At temperatures above "d," it is a *superheated vapor.*

Line HC is thus known by many names – equilibrium, saturated, bubblepoint, dewpoint and vapor pressure. For a pure substance these words all mean the same thing.

At the pressure and temperature represented by HC the system may be all saturated liquid, all saturated vapor or a mixture of vapor and liquid. The exact phase condition of the system depends on the energy level at the P and T involved.

The rectangle "bfghd" illustrates another important phase property that is confirmed experimentally. Suppose we place a liquid in a windowed cell at condition "b" and light it so it is easily visible. We then increase pressure at constant temperature (isothermally). As we proceed toward point "f" the color will begin to fade. At some point (as we blink our eyes) the color disappears completely. The cell now contains what looks like a vapor, but no bubble of vapor was ever seen to form.

At "f" (above the critical) the system is in a fourth phase that cannot be described by the senses. It is usually called *dense phase fluid,* or simply fluid. The word "fluid" refers to anything that will flow and applies equally well to gas and liquid.

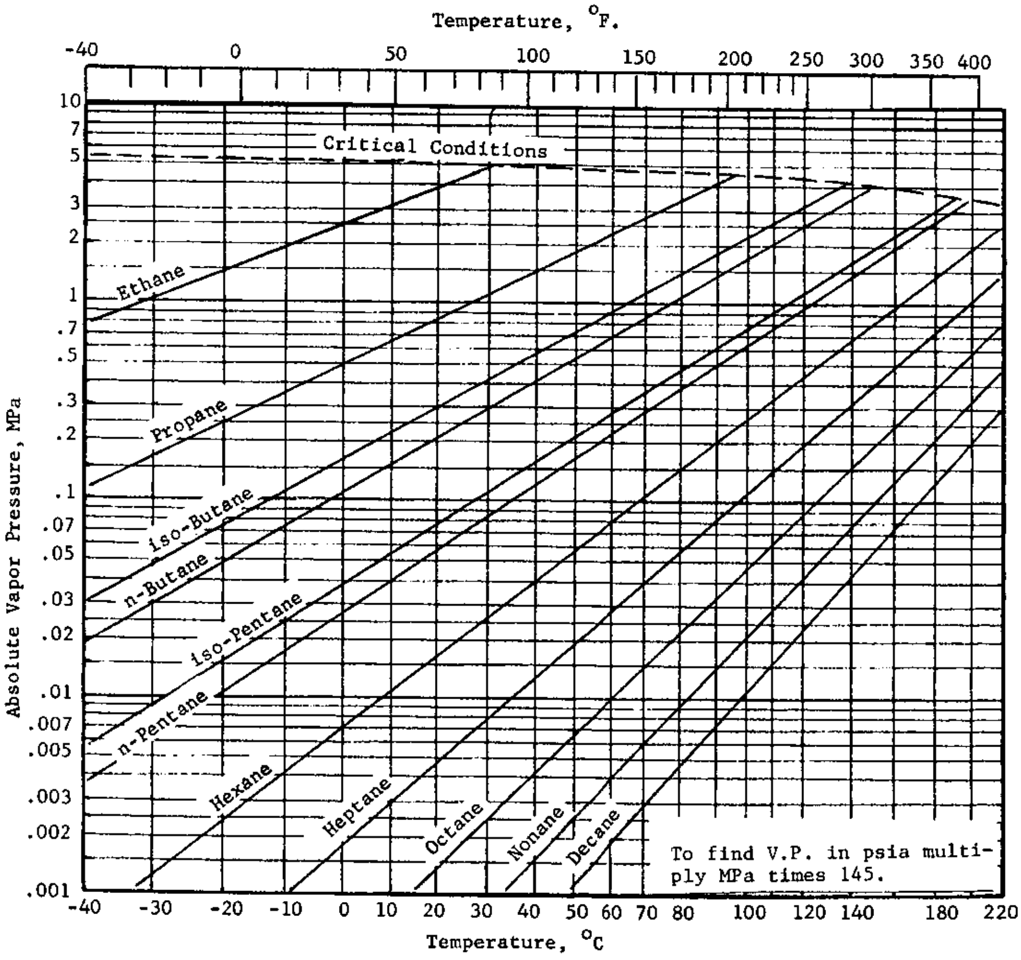
This fluid at "f" looks like a gas but possesses different properties from regular gas found to the right of line HC and below the critical pressure. It is denser than regular gas but is more compressible than a regular liquid. Gas type correlations are used but must be modified to reflect the different behavior patterns of this dense phase fluid.

From "f" one can proceed at constant pressure (isobaric) to "g," reduce pressure isothermally to "h," and then proceed isobarically to "d." One has gone from saturated liquid to saturated vapor without encountering any discernible change of phase.

One could go from "b" and "d" directly by just adding energy to the liquid at constant pressure. In the cell you would observe bubbles of vapor forming and an interface would develop between gas and liquid. As energy input continued the liquid level would fail until the liquid phase disappeared. No temperature change would occur in going from "b" (saturated liquid) to "d" (saturated vapor).

Refer again to Figure 4.1. On the temperature axis you will note t_c. If you follow this line it will go through the critical point "c" and is tangent to the phase envelope HCI. The temperature lines between line HC and CI inside the phase envelope occur at constant pressure. This is a basic characteristic of all such diagrams for pure substances.

Figure 4.3 is a vapor pressure plot for light paraffin hydrocarbons. The lines shown are curve HC in Figure 4.2. They have been straightened artificially by using an odd scale on the abscissa. Figure 4.4 is a corresponding vapor pressure chart for paraffins based on their normal boiling point.$^{(4.1)}$



To find V.P. in psia multiply MPa times 145.

Figure 4.3 Vapor Pressure Chart for Paraffin Hydrocarbons

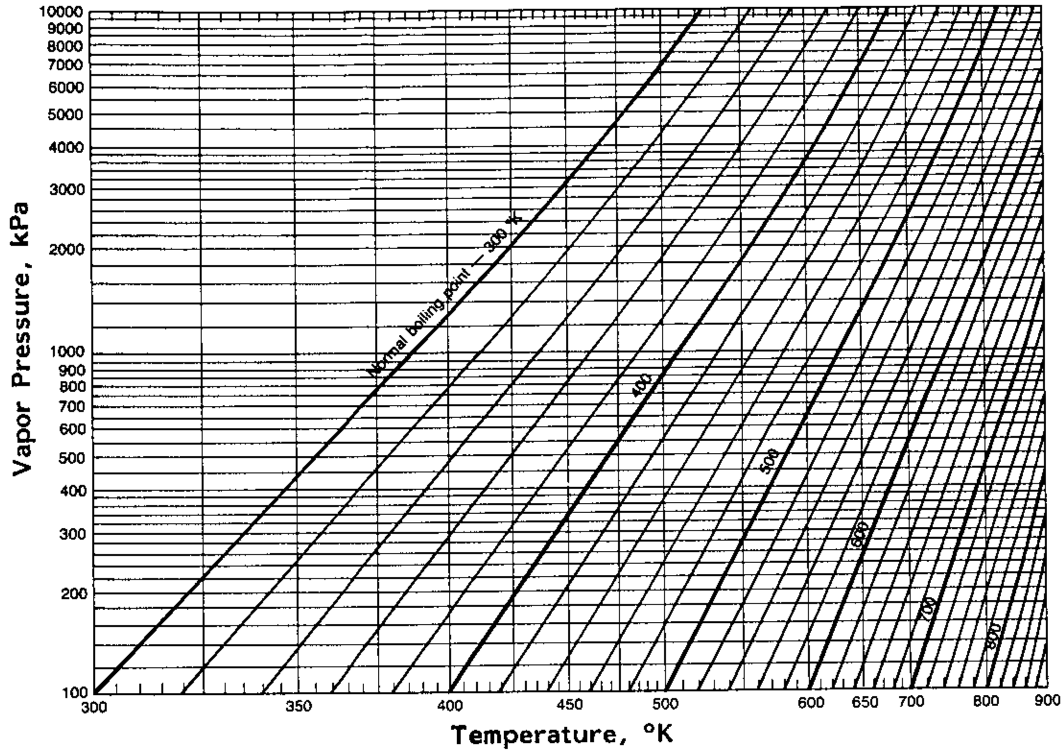


Figure 4.4 **Vapor Pressure Chart for Normal Paraffin Hydrocarbons Based on Normal Boiling Point**

Although the true vapor pressure curve of a pure component must terminate at its critical point, the line often is extrapolated above that point in the calculation of mixture properties. This pseudo value may be used to estimate the contribution of that molecular species to total mixture properties.

MULTI-COMPONENT SYSTEMS

For a multi-component system, another variable must be added to the phase diagram – composition. The location of the lines on a phase diagram depends on composition.

For a pure substance phase envelope HCI is a plane surface parallel to the temperature axis. For a multi-component mixture of substances this phase envelope is not a plane; it has thickness, somewhat like your tongue. Composition is the variable that reflects this thickness. If you replace specific volume in Figure 4.1 by composition and then make a pressure-temperature projection of the solid diagram, you obtain a figure like 4.5. The location of the lines in Figure 4.5 thus varies with composition. This figure is a projection showing only the liquid-vapor portion of the total phase diagram.

There are several terms used to define the location of various points on the phase envelope.

Cricondenbar – maximum pressure at which liquid and vapor may exist (Point N).

Cricondentherm – maximum temperature at which liquid and vapor may coexist in equilibrium (Point M).

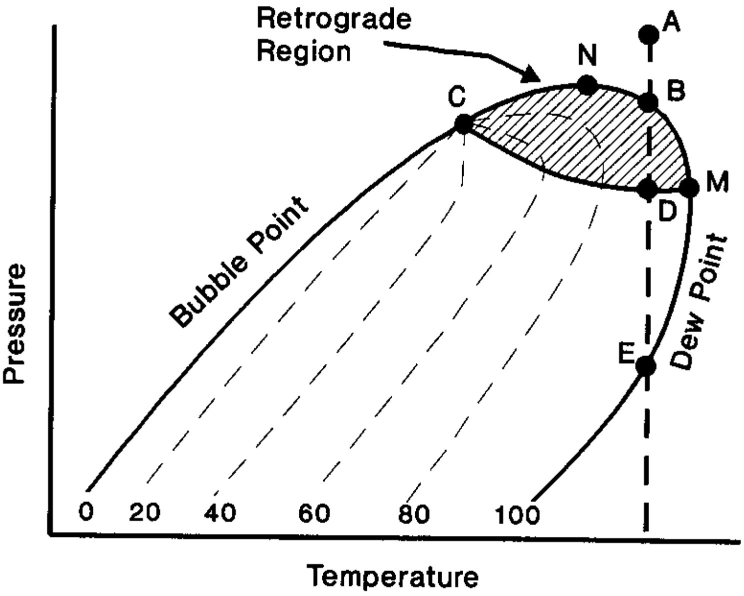


Figure 4.5 **Typical Phase Diagram for a Multi-Component Mixture**

Retrograde Region – that area inside phase envelope where condensation of liquid occurs by lowering pressure or increasing temperature (opposite of normal behavior).

Quality Lines – those lines showing constant percentages which intersect at the critical point (C) and are essentially parallel to the bubblepoint and dewpoint curves. The bubblepoint curve represents 0% vapor and the dewpoint curve 100% vapor.

Line ABDE represents a typical isothermal retrograde condensation process occurring in a condensate reservoir. Point A represents the single phase fluid outside the phase envelope. As pressure is lowered, Point B is reached where condensation begins. As pressure is lowered further, more liquid forms because of the change in the slope of the quality lines. The retrograde area is governed by the inflection points of such lines. As the process continues outside the retrograde area, less and less liquid forms until the dewpoint is reached (Point E). Below E no liquid forms.

In my experience the critical point has always occurred to the left of the cricondenbar for *naturally occurring* hydrocarbon gas mixtures. It is not, however, necessarily in the position shown. It may be further down on the phase curve or closer to the cricondenbar. Location of "C" is most important, for it fixes the shape of the quality lines which in turn governs the vapor-liquid ratio at a given temperature and pressure within the phase envelope.

As shown later in Figure 4.6, the critical point may occur to the right of the cricondenbar. If so, there will be two retrograde regions. But, Figure 4.5 is representative of most systems encountered in production/processing operations.

Note that the pseudocritical values calculated in Chapter 3 are not point "C" on Figure 4.5 except by coincidence. Calculation of the true critical values is discussed later in this chapter.

Effect of Composition

Figure 4.6 for the methane-propane binary system shows the effect of composition on the shape and location of the phase envelope. Two of the lines are the vapor pressure curves for methane and propane, which terminate at their critical point.

Three phase envelopes are shown for three different compositions of methane and propane. This illustrates the fact that the shape and location of the phase envelope depend on composition.

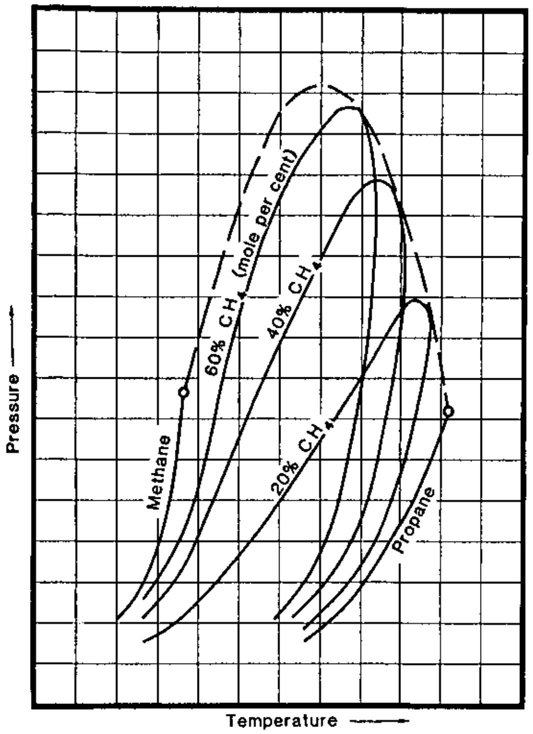


Figure 4.6 **Effect of Composition on Phase Envelope, Methane-Propane Binary**

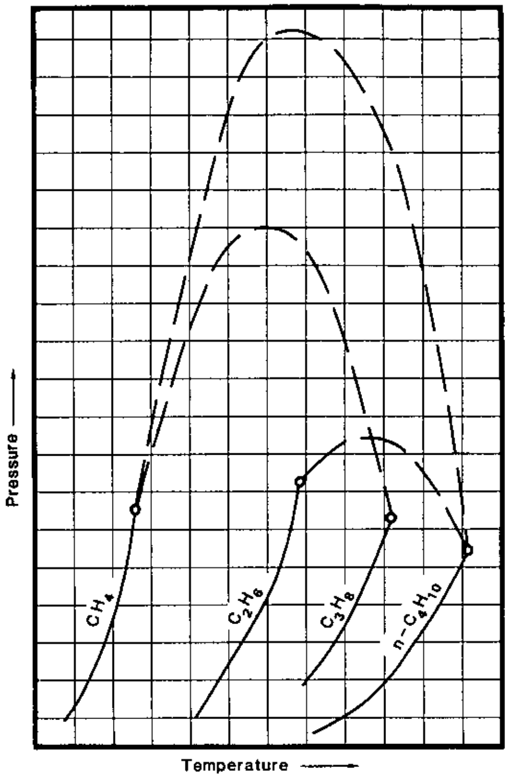


Figure 4.7 **Location of Critical Loci for Several Binary Systems**

The dashed line is the line drawn tangent to all possible phase envelopes or methane-propane binaries at the critical point on each curve. It is called the *critical locus*. It starts at the critical point of methane and ends at the critical point of propane. From the shape of this curve one can deduce that the critical point location on each phase envelope varies with composition.

Figure 4.7 shows the critical loci for a number of common binary mixtures. This behavior of binaries is the basis for the convergence pressure concept used for multi-component mixtures discussed in more detail in Chapter 5. For mixtures of over two components there is more than one critical locus. Thus, convergence pressure is merely a correlation device rather than an exact representation of phase behavior.

Figure 4.6 and 4.7 show that composition has a significant effect on shape and location of the phase envelope. This means that a reliable, representative analysis is important to obtain reliable phase behavior predictions. The accurate prediction of bubble and dewpoint conditions may be critical to design and/or operation. The accuracy of liquid recovery predictions depends on the corresponding quality of the phase data. Chapter 5 will clarify the calculations.

Effect of C_7+ Characterization

As previously discussed, the analysis and/or characterization of the C_6+ or C_7+ fraction in a natural gas mixture is not routine, yet has a significant effect on the shape of the phase envelope. This is illustrated for one gas in Figure 4.8.

Figure 4.8 presents phase envelopes for 4 different characterizations of the C_6+ fraction – C_7, C_9, C_{11} and full characterization based on a distillation analysis.

Special gas chromatographic techniques can identify individual components through about C_8-C_{10}. (See Chapter 1) This is called an extended analysis. If the phase behavior of the gas has a significant effect

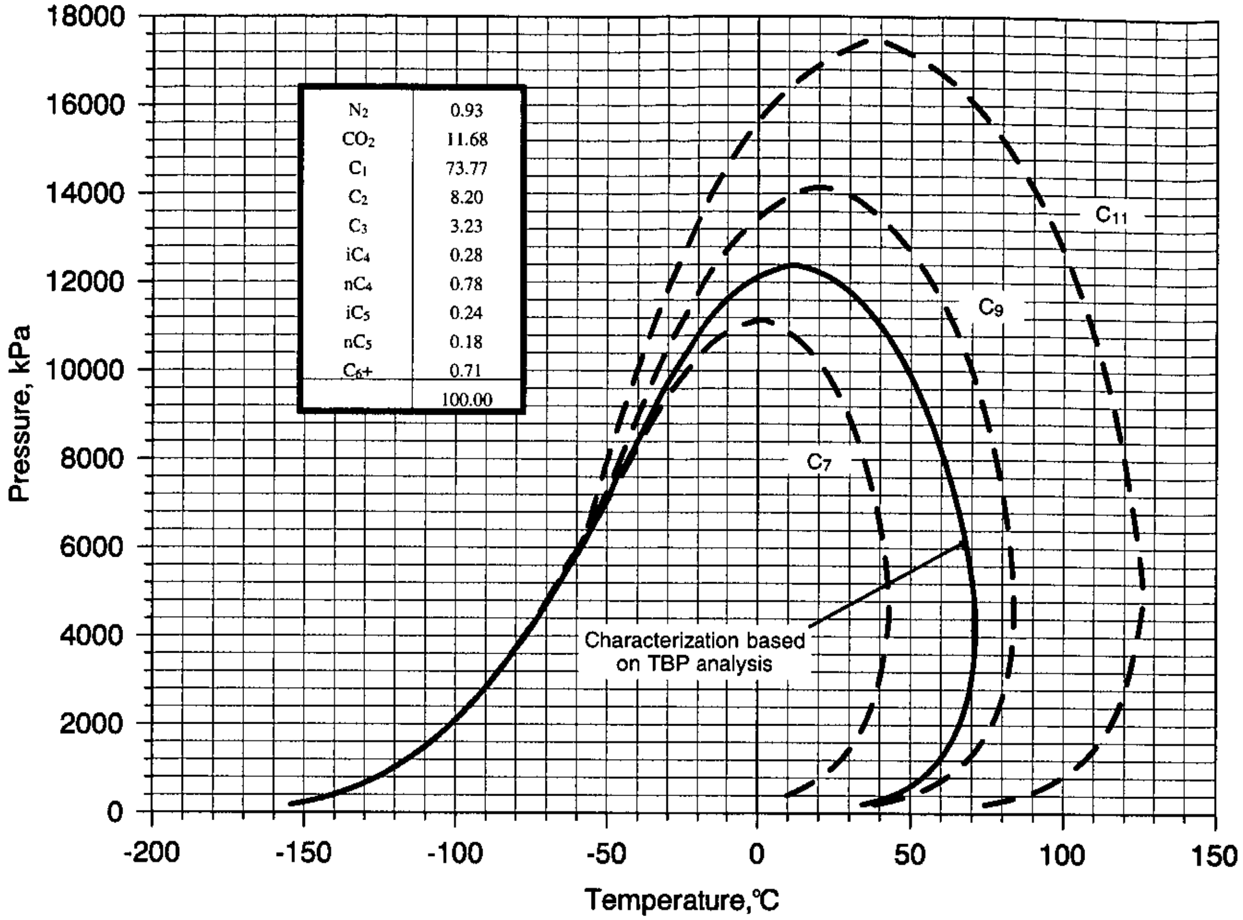


Figure 4.8 Effect of C6+ Characterization on Phase Envelope for Non-Associated Gas

on the system design it is highly recommended that an extended analysis be performed. If an extended analysis is not available, predictive characterization techniques may be used . Several of these are summarized in Chapter 5.

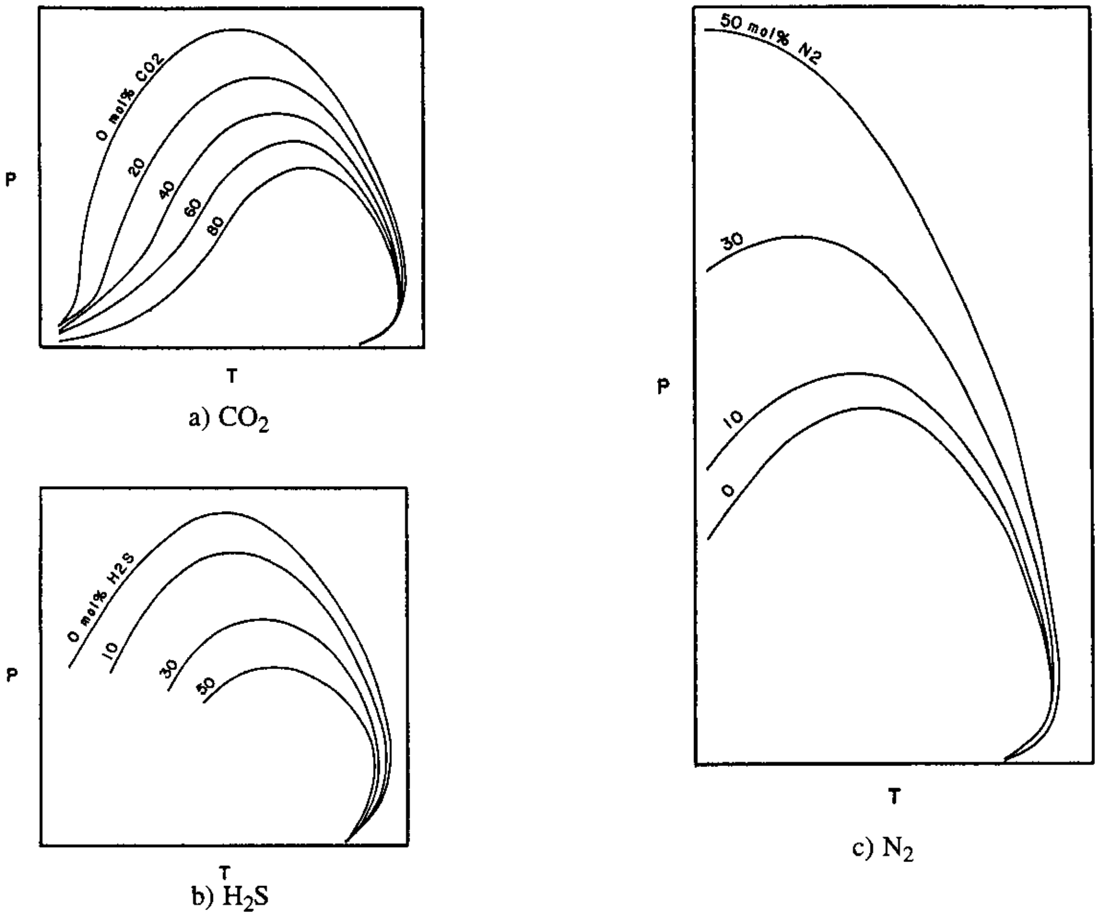
For lean natural gases the C_7+ characterization has a dramatic effect on the dewpoint line. The effect on the location of quality lines is much less significant.

Effect of Impurities

Hydrocarbons are frequently produced with non-hydrocarbon impurities. The most common include water, carbon dioxide, hydrogen sulfide and nitrogen. Since water has a low vapor pressure and is virtually immiscible in the hydrocarbon liquid phase, it does not have a significant effect on the shape of the hydrocarbon phase envelope except at high temperatures and low pressures.

The effect of CO_2, H_2S and N_2 is shown in Figure 4.9. Both CO_2 and H_2S lower the cricondenbar of the mixture. If sufficient quantities of the CO_2 and H_2S components are added to a reservoir fluid and the reservoir pressure is kept above the phase envelope, a single *dense fluid* phase exists. Although the actual mechanism is more complex, it is this solubility that is the primary driving force behind *miscible flood* enhanced oil recovery projects. NGL components such as ethane, propane and butane have a similar effect.

Nitrogen, on the other hand, raises the cricondenbar and decreases miscibility. It is typically used for pressure maintenance.



a) CO_2

b) H_2S

c) N_2

Figure 4.9 **Effect of CO_2, H_2S, and N_2 on Natural Gas Phase Envelope**

APPLICATION OF PHASE ENVELOPES

Proper analysis of many petroleum problems requires knowledge about at least a portion of the phase envelope. Several examples are summarized to illustrate some of these applications.

Reservoir Behavior

The reservoir is the "warehouse" from which we withdraw fluid for production and processing operations. Knowledge about the behavior of this fluid is critical if we are to proceed intelligently.

In Figure 4.10 four different kinds of reservoirs are shown. Points 1, 2, 3 and 4 represent the initial bottom hole pressure. The vertical line from these points represents reservoir pressure decline at constant temperature. The curved line represents pressure and temperature changes that occur in the wellbore. Point "w" represents wellhead conditions.

Reservoir (1) is called a *black oil reservoir.* Its temperature is less than the critical temperature. The reservoir, as shown, is undersaturated. No gas will form in the reservoir until the pressure reaches the bubblepoint, at which point it becomes saturated; as shown, gas would form in the wellbore, however.

Reservoir (2) is usually called a *volatile oil reservoir.* It also occurs to the left of the critical temperature. But, the gas-oil ratio is higher than for a black oil reservoir. The oil may be lighter in color but color is not a reliable indicator of the reservoir type. Once again, no gas forms in the reservoir until the bubblepoint pressure is attained, even though gas forms in the wellbore as shown.

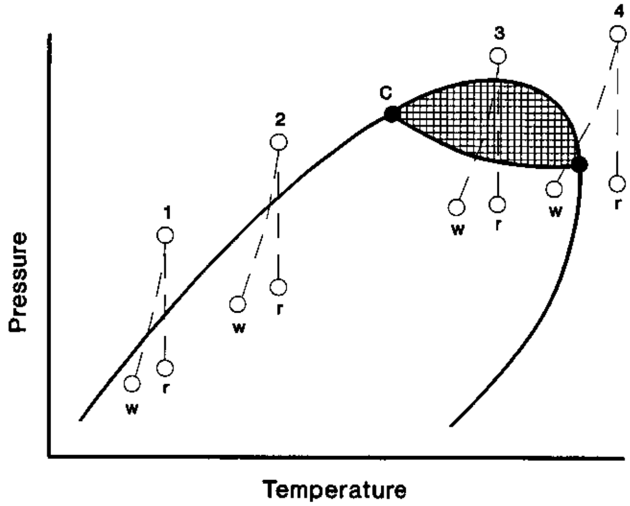


Figure 4.10 Use of Phase Envelopes in Reservoir Prediction

A reservoir whose temperature is between that of the critical and the cricondentherm is a *gas-condensate reservoir.* As it declines in pressure to the dewpoint, liquid forms in the reservoir. Said liquid will not flow to the wellbore until it reaches a critical saturation in the pore space. Most of this liquid will come from the heaviest ends in the dense phase fluid. At the dewpoint, the gas composition in the wellbore might change. The flow rate might also change if non-flowing liquid partially blocks gas flow.

As the pressure declines below the dewpoint, liquid formation increases so long as the pressure is in the retrograde region. Below this region some revaporization occurs.

Reservoir (4) is a true *gas reservoir.* It occurs at a temperature above the cricondentherm. No liquid can form in the reservoir at any pressure. If the wellhead conditions are inside the phase envelope, however, some liquid will form in the wellbore and appear at the surface. So ... once again ... you cannot tell what the reservoir is doing by what you observe at the surface.

In Figure 4.10 the one phase envelope has been used to describe several different types of reservoirs. In actual practice the phase envelopes for each reservoir are different. This is illustrated in Figure 4.11.

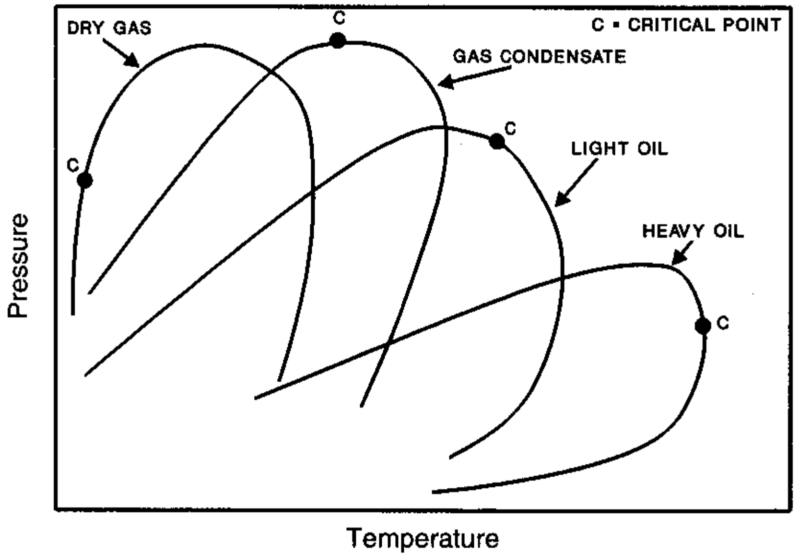


Figure 4.11 Characteristic Phase Envelopes for Four Reservoirs

For a black oil reservoir about all one needs is a section of the bubblepoint curve. For reservoirs that might be volatile oil, gas condensate or gas, what do you need? The upper section of the phase curve, the critical, the cricondenbar and cricondentherm may be necessary. One of the most common mistakes is failure to obtain good reservoir fluid samples and then determine the phase behavior characteristics of said sample. Such failure requires one to make a series of unnecessary "guesses" that may prove to be very expensive.

Pumping Liquids

We often pump commercially pure liquids like ethane and propane at their bubble point. It is important that no vapor form in the pump suction to prevent cavitation. Point A in Figure 4.12 represents the condition in the storage tank (saturated). Point B represents the pressure at the pump inlet due to the head of liquid in the suction piping. Point C represents the pressure inside the pump suction element.

Pressure drop B-C occurs in the pump suction. Point C must be above the saturation curve to prevent cavitation. If the head A-B is not greater than pressure drop B-C, cavitation will occur.

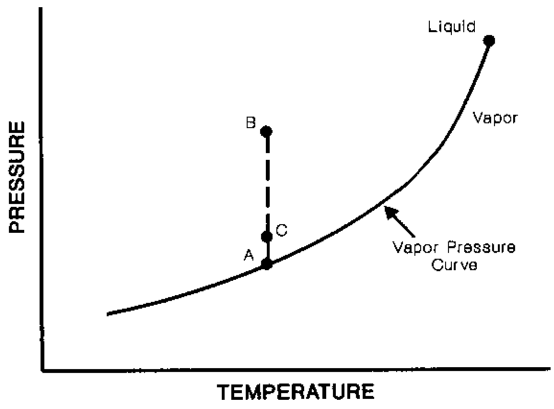


Figure 4.12 Phase Diagrams and Pumping

High Pressure Pipelines

At least a portion of the phase curve must be known in the pipelining of gas.

Consider Figure 4.13. Suppose the solid line represents the phase curve of the gas entering the pipeline and Points I and J represent the inlet and outlet conditions of that line. When line I-J crosses the dewpoint curve, liquid will begin to condense in the line. From that point on, two-phase flow exists.

Liquid can be prevented by removing enough of the heavy ends to change the shape and position of the phase curve. Suppose in our example that the composition is changed enough so that the dashed line results. No liquid will form in the pipeline because the dewpoint curve is not crossed. Any liquid present would be due to separator liquid carryover or a composition different from the one used to construct the phase curve.

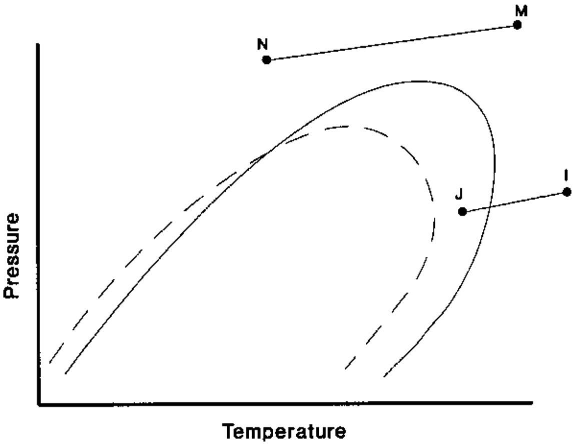


Figure 4.13 High Pressure Pipelines

Line M-N is a different situation. This represents operation of a pipeline in the *dense fluid* region. This is frequently done to avoid liquid condensation in the line and the associated problems of designing and operating a two-phase system. There are two very pertinent questions: (1) Will line M-N intersect the phase envelope and liquid form? (2) Will Point N end up below the cricondenbar, to the left of the phase curve, so that the line will have more nearly liquid characteristics than those of a dense phase fluid at its discharge end?

To answer these questions requires a good analysis and a reliable phase curve (preferably from a laboratory determination). If separation has occurred between the wellhead and the line, the analysis entering the line must be used to construct the phase curve. The phase curve is very sensitive to composition. One only has to look at Figure 4.8 to be reminded of this fact.

Refrigeration Processes

Refrigeration is frequently used to cool a gas in order to meet a hydrocarbon dewpoint specification. When the refrigeration takes place at high pressure, it is very important to have an accurate description of the phase behavior. Figure 4.14 shows the phase envelope for a lean gas typical of that found in the southern North Sea or Gulf of Mexico. The cricondentherm of this gas is 30°C [86°F] and occurs at a pressure of 3000 kPa [435 psia].

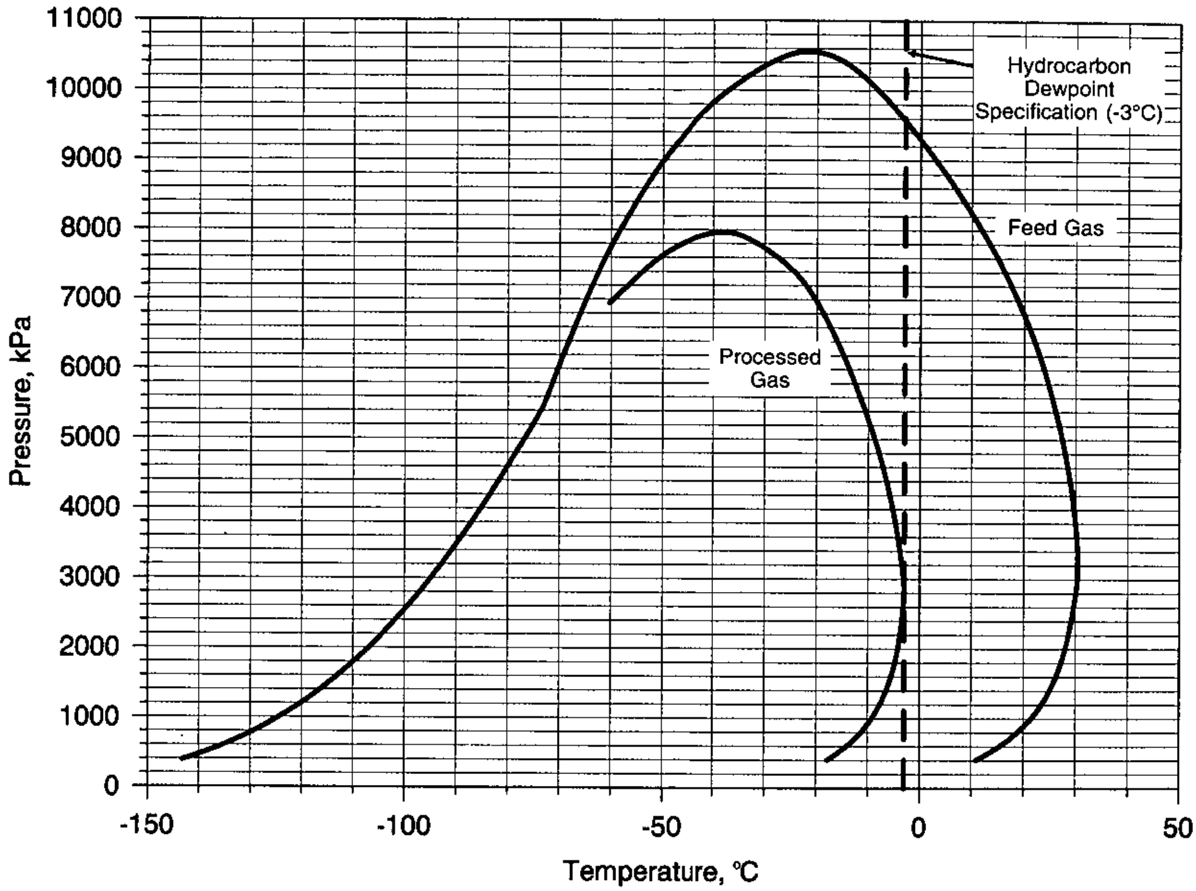


Figure 4.14 Effect of Pressure on Dewpoint Control Processing Temperatures

Assume that this gas is to be processed in a mechanical refrigeration plant to meet a hydrocarbon dewpoint specification of –3°C [27°F]. If the processing pressure is 3000 kPa [435 psia], the gas must be cooled to at least –3°C [27°F] to meet the dewpoint specification. However, if the processing pressure is 7500 kPa [1088 psia], the gas must now be cooled to –25°C [–13°F]. The following shows the processing temperature required to meet the dewpoint specification at various process temperatures.

P, kPa	T, °C
3000	–3.0
5000	–7.2
7000	–19.4
7500	–25.1
7900	–34.0

Above about 8000 kPa [1160 psia] this gas could not be processed in a mechanical refrigeration plant to meet the hydrocarbon dewpoint specification. In order to meet the specification an expansion process (valve or expander) or an adsorption process must be used.

Operation Near the Critical Region

In some cases, having an accurate phase envelope is not good enough; you also need a reliable, true critical point. The pseudocritical found from PVT models are seldom reliable. This may be illustrated by the quality lines discussed previously. If, in Figure 4.15, the critical is at C_1 the 20% and 80% quality lines might be as shown. Suppose, instead, that the true critical is at C_2. Notice the possible change in the location of the same quality lines. If the operating pressure is much less than C_2, the error in the amount of liquid predicted will be relatively small.

However, if the operating pressure falls between C_2 and C_1, the error can be significant – having a profound effect on system design and/or operation. In general, operation near the critical point should be avoided. Common sense dictates that if the system is difficult to model it will be difficulat to operate and control. If near-critical operation is unavoidable, the process design should reflect the uncretainties involved. the most common examples of potential near-critical operation are low temperature expander plants and de-ethanizers.

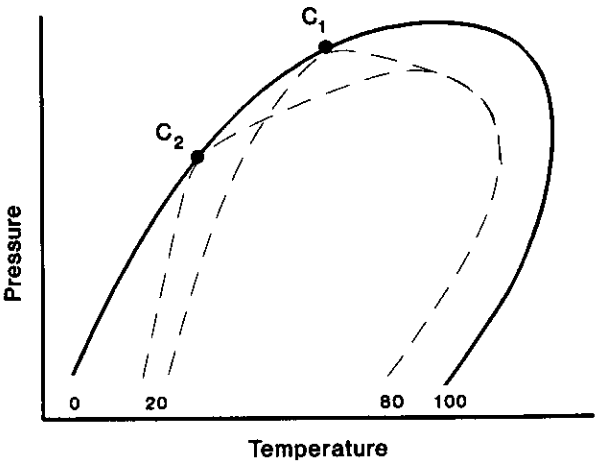


Figure 4.15 Refrigeration Processes

Practical Suggestions

The above are merely examples of typical problems that have arisen. One could cite many others. Failure to handle phase behavior in a workmanlike, professional manner has proven, and will continue to prove, very expensive for the petroleum industry. The cost of obtaining good data is usually trivial compared to the economic benefits obtained.

There are some good guidelines that should be followed:

1. Obtain good samples using experienced samplers.
2. Handle the samples carefully.
3. Analyze the samples in a proven laboratory.
4. Develop all, or that portion needed, of the phase curve. Above about 14 MPa [2000 psi] calculated points are suspect no matter how big the computer or how complex the program. The results should be regarded as estimates only. If an estimate is good enough, fine. If not, go to a reliable laboratory. In many cases a few laboratory points supplemented by calculations will suffice.

GIBBS PHASE RULE

One of the applications of phase behavior is to control the state of the system. In many applications the phase rule is useful. It is written:

$$F + P = C + 2 \tag{4.1}$$

Where:
F = number of degrees of freedom which a system possesses
P = number of phases in the system
C = number of components or chemical compounds in the system

The degrees of freedom refer to the variables necessary to fix the phase state of the system. One or more of the following variables are commonly used: pressure, temperature, volume, and total or partial composition of the one or more phases present.

Consider the single component system shown in Figure 4.2. The number of components equals one. Therefore, F + P = 3. Along line HC (Figure 4.2), if we have two phases in equilibrium, P = 2, and F = 1. Therefore, specification of either pressure or temperature would be sufficient to fix the system state. At point "h" only one phase is present (P = 1) and F = 2. Therefore, knowledge of both pressure and temperature would be necessary to fix the state of the system (exact location of point "h" on phase diagram).

At the triple point (pure component) there are three phases present, so P = 3. Since C = 1, F = 0 from Equation 4.1. This means that once you say "triple point" you have completely defined the system since there is only one triple point.

The number of phases present is not necessarily limited to three. There can be only one gas phase but there can be more than one liquid and solid phase.

Suppose you have a system containing gas, liquid hydrocarbon, liquid water, ice and hydrate. The two liquid phases are *immiscible* and possess different properties. Ice and hydrate are both solids and rather white in color, but they possess different physical characteristics and properties. So ... how many phases are present? Five!

A phase is any entity within the system that contains separately identifiable behavior and properties.

A miscible mixture is made up of molecules possessing different properties, but we cannot measure these separately in the mixture state. So ... this mixture is only one phase.

The number of components used in Equation 4.1 depends on how much we know about the system, or need to know, for the application involved. If we are dealing with an equation or correlation where composition is not important, we can handle any phase as one component; simply as gas, liquid or solid. If composition is important we can handle each component in the system as a phase rule component.

Example 4.1: Suppose we have a system for which nine components are analyzed ($C_1 - C_7+$); "C" in Equation 4.1 will be 9. Suppose further that we have a two phase system (liquid-vapor).

$$F + 2 = 9 + 2 \quad , \quad F = 9$$

We must fix nine variables to fix the state of the system. How many independent variables do we have to work with? We have eight independent component compositions (the ninth is fixed by difference and therefore is not independent), pressure and temperature, to name the obvious ones. If we fix any nine of these ten, the state of the system is fixed.

The Gibbs Phase Rule is a factor in the design of controls for systems. A common problem is excess control. When there are more controls than needed they fight each other and system stability suffers. Suppose one is using pure propane in a refrigeration chiller. Line HC in Figure 4.2 represents the P-T relationship. From Equation 4.1, F = 1. So, control of pressure fixes a point on line HC which in turn fixes temperature; fixing temperature through control likewise fixes pressure. Control of both P and T is not needed.

Degrees of freedom (F) will be some number greater than unity everywhere on a phase curve except along the phase boundary lines. Unless the control system is compatible with the phase behavior and Gibbs Phase Rule, it will not be entirely satisfactory.

PREDICTION OF PHASE ENVELOPE

The location of the bubblepoint and dewpoint lines may be calculated using the methods of Chapter 5. For most naturally occurring systems above about 14 MPa [2000 psia], the validity of the standard calculation becomes questionable. If the location of the curve at high pressure is very important – and if a reliable fluid sample is obtainable – a laboratory determination of at least selected parts of this curve is recommended.

However, there are circumstances where the calculation of all or part of the upper phase curve may be satisfactory. References 4.2 on, at the end of this chapter, provide further details.

Estimation of the critical, cricondentherm and cricondenbar points is particularly important.

Cricondentherm and Cricondenbar T and P

Grieves and Thodos have prepared correlations for these points based on 123 binary and 15 multi-component mixtures.*(4.3,4.4)* For the systems tested, the agreement between predicted and measured values was very satisfactory. The prediction of temperature is better than pressure. The maximum is less than 5%, whereas the maximum pressure deviation was around 13%.

These correlations are useful in conjunction with vapor-liquid equilibrium (VLE) calculations and critical point predictions. Usually in all of these it may be possible to construct a phase curve (or portion thereof) that is useful for at least planning purposes. This may be the case in the early stages of development when good fluid samples are not yet available. Sometimes the only samples are from drill–stem tests or nonproduced exploratory wells, which taints the validity of the laboratory results. Some calculated cricondentherm and cricondenbar numbers may prove helpful, for they denote the maximum T and P of any phase curve.

Critical Pressure and Temperature

The prediction of the location of the critical point is difficult. The most reliable way remains a laboratory study of a reliable sample. But, there are many circumstances where a calculated number may be suitable.

Based on an empirical fit of 25 natural gas systems possessing a molecular weight less than 30, the following equation has been developed.

$$T_c/T_c' = 1.0 + (0.03)(M\text{ gas} - 16) \tag{4.2}$$

Where: T_c = actual critical temperature
T_c' = pseudocritical temperature from Kay's Rule (Chapter 3)

For the samples tested the error was less than plus or minus 5%.

A more accurate method*(4.5)* uses the equation

$$T_{c_{mix}} = \sum \varphi_i T_{c_i} \tag{4.3}$$

Where: $T_{c_{mix}}$ = critical temperature of mixture
T_{c_i} = critical temperature of each component
φ_i = critical volume fraction as defined in Equation 4.4

$$\varphi_i = \frac{y_i \nu_{c_i}}{\sum y_i \nu_{c_i}} \tag{4.4}$$

Where: y_i = mol fraction
ν_{c_i} = critical volume of each component

A similar method(4.6) has been proposed which employs third parameters and more complex mixing rules, but the increased accuracy is marginal.

No simple, satisfactory correlation has been found for critical pressure. References 4.6, 4.7, and 4.8 outline some methods available.

One problem is prediction of the critical values for the hydrocarbon fractions heavier than hexane shown in analyses. Correlations based on the mean boiling point of the fraction, as found from distillation test, have proven useful.(4.2, 4.9)

Simon and Yarborough(4.10) present a correlation for predicting critical pressure based on 14 gas-solvent reservoir systems using composition and the heptanes plus fraction molecular weight and characterization factor as parameters. A total of 37 other systems shown in the literature were tested. The average deviation was about 5%.

Whitson(4.11) has proposed a series of generalized equations for property prediction, including critical values for heavier fractions such as heptanes plus. The following equations were proposed that used earlier data.(4.12)

The value of T_b in Equations 4.5 and 4.6 may be found from a distillation analysis or from a mixing rule.

$$P_c = a(T_b)^b \, (\gamma)^c \tag{4.5}$$

		Metric	English
Where:	T_b = average boiling point	K	°R
	γ = relative density	–	–
	P_c = critical pressure	kPa	psia
	a = empirical constant	5.53×10^9	3.12×10^9
	b = empirical constant	–2.31	
	c = empirical constant	2.32	

Equation 4.5 applies for any liquid having a boiling point less than 472 K [850°R].

$$T_c = a(T_b)^b \, (\gamma)^c \tag{4.6}$$

		Metric	English
Where:	T_c = critical temperature	K	°R
	a = empirical constant	19.1	24.3
	b = empirical constant	0.59	
	c = empirical constant	0.36	
	γ = relative density	–	

It is recommended that references like those cited immediately above be used with caution and only by those very familiar with phase behavior. They are shown to illustrate the type of simple, manual correlations available.

Equations of state provide the most accurate and consistent methods of predicting critical properties. However, even these methods provide an approximation only.

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NOTES:

5

VAPOR-LIQUID EQUILIBRIUM BEHAVIOR

Many production/processing operations involve control of the phases present. Selected components must be vaporized or condensed to achieve specified goals. For a given mixture fluid composition the phase envelope defines the vapor, liquid and two-phase regions. Determination of the bubble and dewpoint curves fixes the extent of the two-phase region. Within this region it is necessary to predict vapor-liquid ratio and the composition of both phases. The latter involves the concept of equilibrium.

"Equilibrium" is the word commonly used to describe a state wherein the passage of further time will have no effect on system behavior. In a flow system, the rate is independent of time. In a phase behavior context, the composition of the vapor and liquid in contact will remain constant with time.

It takes time for two phases to reach equilibrium. In a typical hydrocarbon system up to several minutes of contact is required. Since a well-designed separator will have a vapor-liquid contact time of this magnitude it is assumed normally that a separator should be calculated as an equilibrium process.

In absorption and fractionation calculations no true equilibrium exists because the vapor and liquid are not in contact long enough to achieve it. The assumption of equilibrium (made for convenience) must be corrected by an efficiency factor.

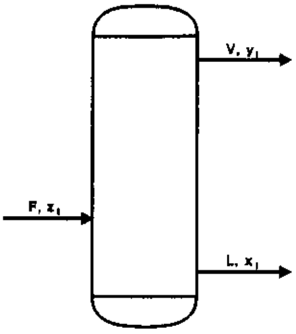
The concept of equilibrium in a two phase system does not mean that the system is static. There is still a continuous transfer of molecules between phases. The rate of vaporization and condensation of all molecular species is equal. Thus, no net change in composition occurs.

EQUILIBRIUM VAPORIZATION RATIO

It is convenient to represent equilibrium with an *equilibrium vaporization ratio*, K. This is defined by

$$K_i = \frac{y_i}{x_i}$$

Values of z_i, x_i and y_i are specified or found from sampling and the resultant analyses. Values of F, V and L are specified or measured (for an existing system). The detailed procedure for using this balance will be discussed in a later section.

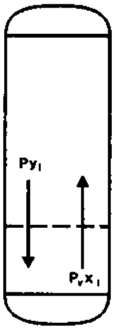


This K_i may be incorporated into a material balance around a separator, where:

z_i = mol fraction of any component in total feed stream to separation vessel
y_i = mol fraction of any component in the vapor phase
x_i = mol fraction of any component in the liquid phase
K_i = equilibrium vaporization ratio (equilibrium constant) = y_i/x_i
F = total mols of feed
V = total mols of vapor
L = total mols of liquid

Partial Pressure Concept

At pressures up to about 400 kPa [60 psia], where ideal gas concepts apply, vapor pressure may be used to find equilibrium behavior.



At equilibrium, the rate of vaporization of all components must equal the rate of condensation so that no change occurs in vapor or liquid composition. For this to occur, the driving force in both directions must be equal. This driving force may be represented by a partial pressure (pp).

$$\text{pp (vapor phase)} = P\,y_i$$
$$\text{pp (liquid phase)} = P_{v_i}\,x_i$$

Where: P_{v_i} = vapor pressure of any component at the P and T of separation

At equilibrium,
$$P\,y_i = P_{v_i}\,x_i\ , \quad \text{or} \quad K_i = \frac{P_{v_i}}{P} \qquad (5.1)$$

Equation 5.1 applies anywhere within the phase envelope. It does not apply for the liquid water phase in a hydrocarbon system since water is essentially immiscible with liquid hydrocarbons. (Chapter 6 will discuss this in detail.) Equilibrium calculations on the hydrocarbon portion of the system are not influenced to a measurable degree by the presence of water except at low pressures and high temperatures.

Equation 5.1 is also the best and most convenient way to find K at pressure approaching atmospheric. P_v may be found from vapor pressure data and simply divided by total absolute pressure P. Also, the values obtained are more reliable than other methods of obtaining K discussed in later sections, since these oftentimes fail to converge properly at low pressures.

Fugacity Derived K Values

The term fugacity (f) is a conceptual term related to free energy and other basic thermodynamic concepts. It has the units of pressure. Since it is one way to correct for nonideal behavior, and may be related to compressibility factor Z, it is sometimes referred to as a "corrected pressure." Some correlations are developed using fugacity coefficient (f/P), a convenient dimensionless term.

The partial pressure concept can be extended to about 3.5 MPa [500 psia] by using fugacity as a representation of driving force. At a given equilibrium pressure and temperature,

$$K_i = \frac{y_i}{x_i} = \frac{f^o_{i_L}}{f^o_{i_V}} \qquad (5.2)$$

Where: $f^o_{i_L}$ = fugacity of component "i" in the pure liquid state
$f^o_{i_V}$ = fugacity of component "i" in the pure vapor state

According to the law of ideal solutions (not to be confused with the ideal gas laws)

$$f_{i_L} = f^{o}_{i_L} x_i \quad \text{and} \quad f_{i_V} = f^{o}_{i_V} y_i \tag{5.3}$$

Where: f_{i_L} = fugacity of component "i" in the liquid mixture

f_{i_V} = fugacity of component "i" in the vapor mixture

At equilibrium, $f_{iV} = f_{iL}$, because of the necessary free energy relationship. From Equation 5.3, $f^{o}_{iL}x_i = f^{o}_{iV}y_i$, and $y_i/x_i = f^{o}_{iL}/f^{o}_{iV} = K_i$.

Computer Solutions for K

Today the majority of the calculations utilizing K are made on computers. In general, determination of K values follows one of the two methods below:

1. Equation of State Method

As previously indicated, fugacity is a thermodynamic property which is related mathematically to P, V & T. For convenience, fugacity coefficients, φ, are frequently used where $\varphi_i = f_i/y_iP$

@ equilibrium $\qquad f_{i_V} = f_{i_L} \quad \text{so} \quad K_i = \dfrac{\varphi_{i_L}}{\varphi_{i_V}}$

Fugacity coefficients can be calculated from Equation 5.4 where the P, V, T relationships are provided by an Equation of State (PR, SRK, BWRS, etc.).

$$\ln \varphi_i = \frac{1}{RT}\int_V^\infty \left[\left(\frac{\delta P}{\delta n}\right)_{T,V,n_j} - \frac{RT}{V}\right] dV - \ln Z \tag{5.4}$$

This is the most common method of determining K values for hydrocarbon mixtures. Why don't different programs yield the same K values?

First of all, the calculation involves solving P-V-T equations of state, which in turn requires the use of combination rules. As with Z, H, S and other quantities, there is no one correlation that fits all systems equally well. A given approach is the result of fitting data for a specified system empirically.

So long as the molecules involved are similar, the problem of fitting is relatively routine. Predicted values for the normal paraffins ethane through decane are usually quite reliable. Methane poses a problem because of its high vapor pressure and unusual behavior in the liquid phase. Nitrogen poses a similar problem. One can likewise correlate values for the aromatics. Hydrogen sulfide, water and carbon dioxide act differently from hydrocarbons because of their polarity.

If one has a mixture containing many, or all, of the above components, the behavior is rather complex. As noted in Chapter 3 for other properties, special correlations or computer subroutines are necessary for some mixtures.

2. Activity Coefficients

Equation of state methods generally do a reasonably good job of determining vapor phase fugacities. The liquid phase is much more difficult, especially when the liquid contains dissimilar molecules. Polar molecules like H_2S, H_2O, alcohols and glycol are not easily handled. In situations like these the liquid phase fugacity is calculated from Equation 5.5.

$$f_{i_L} = f^o_{i_L} x_i \gamma_{i_L} \qquad (5.5)$$

where the activity coefficient, γ_i, is an empirically determined correction factor.

The activity coefficient (γ_{i_L}) approaches unity as the concentration of any component approaches 100%. It is expressible by the Margules, van Laar, Wilson or Renon type equations. Standard thermodynamics may be used to relate the effect of temperature and pressure on activity coefficient. The effect of composition is not so easy. The Hildebrand equations may be used to estimate activity coefficients with satisfactory accuracy for nonpolar mixtures. The K value is then determined from Equation 5.6.

$$K_i = \frac{y_i}{x_i} = \frac{f_{i_L} \gamma_i}{P \varphi_{i_V}} \qquad (5.6)$$

The Chao Seader method is an example of a K value correlation which utilizes activity coefficients.

Stated simply, there is no one, magic correlation or program that works equally well for all mixtures. The choice of program must be made by one who has experience with the use of a given calculation and knows the kind of fit to be expected so adjustments can be made. This is not an area for an "amateur" even though he knows how to use the computer and can obtain a printout. Most problems resulting from the use of K values result more from the persons developing and using them than from the weaknesses of the calculations themselves.

A K value is a function of pressure, temperature and composition. Of the two phases present, liquid composition is the most critical. One difference between correlations is the method chosen to find the coefficients used to characterize the liquid phase.

Those who specialize in the K value area generally have a favorite method (often theirs) which they promote with all the fervor of a religious zealot. The net result is a somewhat chaotic series of claims that might confuse more than help. There is only one true criterion for evaluating K values. How well do they check the performance of equipment designed from them? Once the equipment is operating, get some actual values. Compare them with the correlation used. Use this information to modify the correlation or to develop alternatives. A sound set of values used in this manner will produce useful, consistent results.

K Value Plots

The K value of any component depends on pressure, temperature and the composition of the total fluid involved. Traditionally, many plots of log K versus P and T have been proposed. The effect of composition may be ignored by making the plot for a specific group of components. Up to a pressure of about 6.9 MPa [1000 psia] the effect of composition on K usually is not critical although it may affect methane significantly.

Two sets of K values are summarized in two appendixes at the end of this chapter. Appendix 5A is a series of computer-generated charts using SRK equations. The values shown are useful particularly for the calculation of systems discussed in this book wherein liquid is being condensed from gas systems.(5.1)

Appendix 5B is based on data obtained from field tests and correlations on oil-gas separators.(5.2) The data set was based on over 3000 values. This correlation has been used often for oil separation calculations.

The values in Appendix 5A will be used in all examples in this book to achieve internal consistency. These values also are a printout of the K value program developed for use with this book.

Convergence Pressure Concept

Almost all current applications of K values involve calculations made on a programmable calculator or a computer. But, charts still are available which use convergence pressure (P_k) as a parameter to represent compositional effects. Thus, this brief summary of the concept involved.

Figure 4.7 contained a series of lines which were the loci of the critical points of all possible analyses of a given binary mixture. The location of these lines depended on the components making up this binary, since they were drawn between the critical points of said components. The concept of convergence pressure was based on this binary behavior. References 5.3-5.8 present methods for using the convergence pressure concept.

Figure 5.1 is a plot of log K vs. log P (at constant temperature) for four different mixtures where methane is the lightest component and normal decane the heaviest.(5.3) The curve shape is typical of K value plots. The curves for propane and heavier have a negative slope with increasing pressure until they reach an inflection point after which they have a steep positive slope.

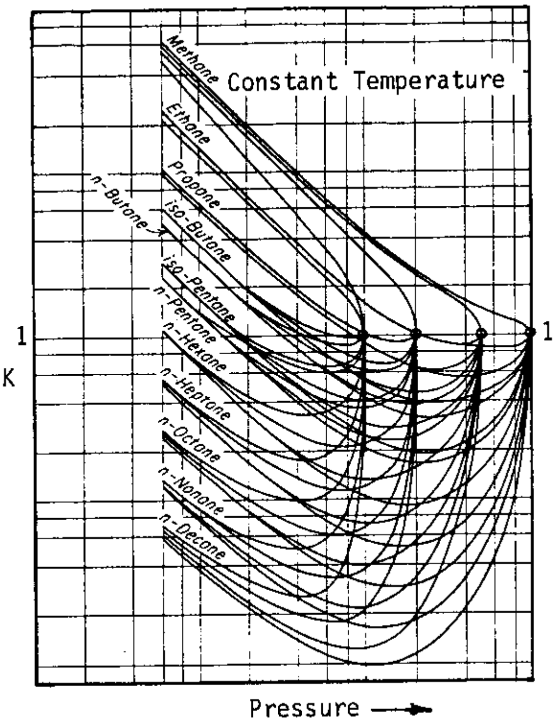


Figure 5.1 Effect of Composition on K Values

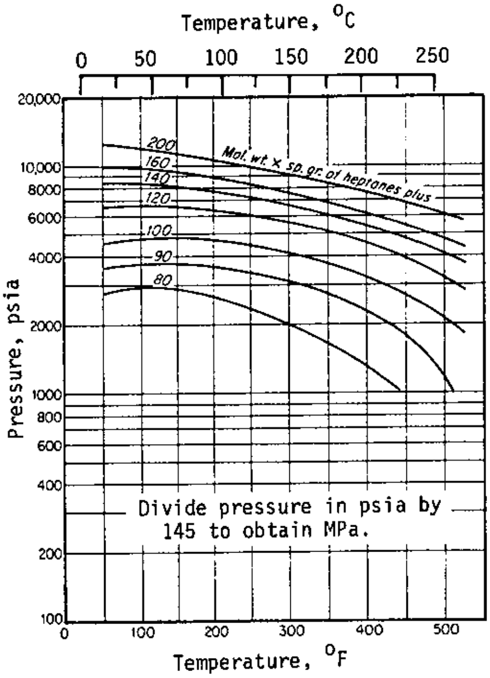
As shown, the position of the curves changes with composition. For any one composition the curves converge at K = 1.0. The pressure of this convergence is called *convergence pressure*. As shown in Figure 5.1, it varies with composition. If the temperature at which the curves were plotted was the critical temperature, the convergence point would be the critical pressure. At critical conditions K = 1.0 for all components. However, at any other temperature convergence pressure is not on the phase curve; it merely is a correlating number.

Figure 5.2 is a convenient way to estimate convergence pressure for a hydrocarbon mixture containing methane as the lightest component. It is included for the convenience of those readers who might have occasion to estimate such a number for a manual calculation.

Notice in Figure 5.1 that the slope of the K value curves is very steep to the right of the slope change inflection point. In this section it is difficult (if not impractical) to read accurate values of K. For this reason, this type of correlation should not be used where P is within about 7 MPa [1000 psi] of the convergence pressure. Above a P of about 14 MPa [2000 psi] such correlations have limited utility.

K Values for Heaviest Fraction

When one is using a K value plot or table of values, an extrapolation may be necessary to find a K for the heaviest fraction. As shown in Appendix 5B, the mean average boiling point of the fraction may be used. This fraction may vary from pentanes plus to nonane plus depending on the system.



Divide pressure in psia by 145 to obtain MPa.

Figure 5.2 Estimation of Convergence Pressure from Heptanes Plus

It usually is satisfactory to use the heptanes plus fraction for gases separated from crude oil at, or near, ambient temperatures. With the molecular weight and specific gravity of this fraction one can predict gas behavior very satisfactorily.

For gas condensate and crude oil systems, the nonanes plus fraction is very desirable for the heaviest fraction is usually a significant part of the total stream. In specifying the analysis of a sample from these streams, you should require an analysis through not less than nonanes plus and a boiling point curve on this heaviest fraction.

What do you do if you already have an analysis for a crude oil system and it goes only through heptanes plus? Several techniques have been proposed for creating a pseudo analysis of this fraction to better predict K values.(5.10-5.14) The simplest approach assumes a logarithmic probability distribution of heavy molecules and that nothing but normal paraffins are present.

For systems handling reservoir produced fluids, compounds other than paraffins are present. A normal chromatographic analysis does not reflect this fact. Most of the aromatics, napthenes and the like will be shown as n-hexane or heavier. So ... in many cases we also recommend more than an ordinary chromatographic analysis be run to identify the type and quantity of nonparaffins. This is important knowledge.

For many calculations it is sufficient to predict a single value of K for the heptanes plus fraction. The five approaches outlined are in the order of my personal preference.

1. **Log K vs. T_c^2**

   We have found that this plot is linear for all hydrocarbon K data studied to date. There are some semiscientific reasons to support this result. What is more important, this relationship serves two practical purposes: (1) checking K values for internal consistency and (2) extrapolating pure component K values to obtain the K of the heaviest fraction.(5.9)

   Figure 5.3 is an example of the chart we use. The abscissa is really T_c^2 but has been drawn to reflect the component and the molecular weight of the heaviest fraction. It is used as follows:

   a. Plot K values for each pure component at the P and T of the equilibrium.
   b. Draw the best straight line through the points for propane through hexane components.
   c. If any K value points are not on the line, use the line values for any calculation.
   d. Find the K value of the C_7+ fraction from its molecular weight and the straight line extrapolation.

   If the K values come from a computer program the values may be rectified into a straight line by regression analysis since the equation of the straight line is $\log K = a + bT_c^2$.

2. **Volatility Exponent**

   This is a correlation expressed by the equation

$$K\,(C_7+) = \frac{K_7}{(K_2/K_7)^b} \tag{5.7}$$

Where:
b = exponent found from Figure 5.4
K_2 = K of ethane
K_7 = K of heptane

Figure 5.4 requires a knowledge of the atmospheric boiling point of the heptanes plus fraction. If not known from other sources, use Figure 5.5 to find it. Values of K for ethane and heptane are those used in the calculation. This approach has been used for a long time and is reliable for most natural gas streams.

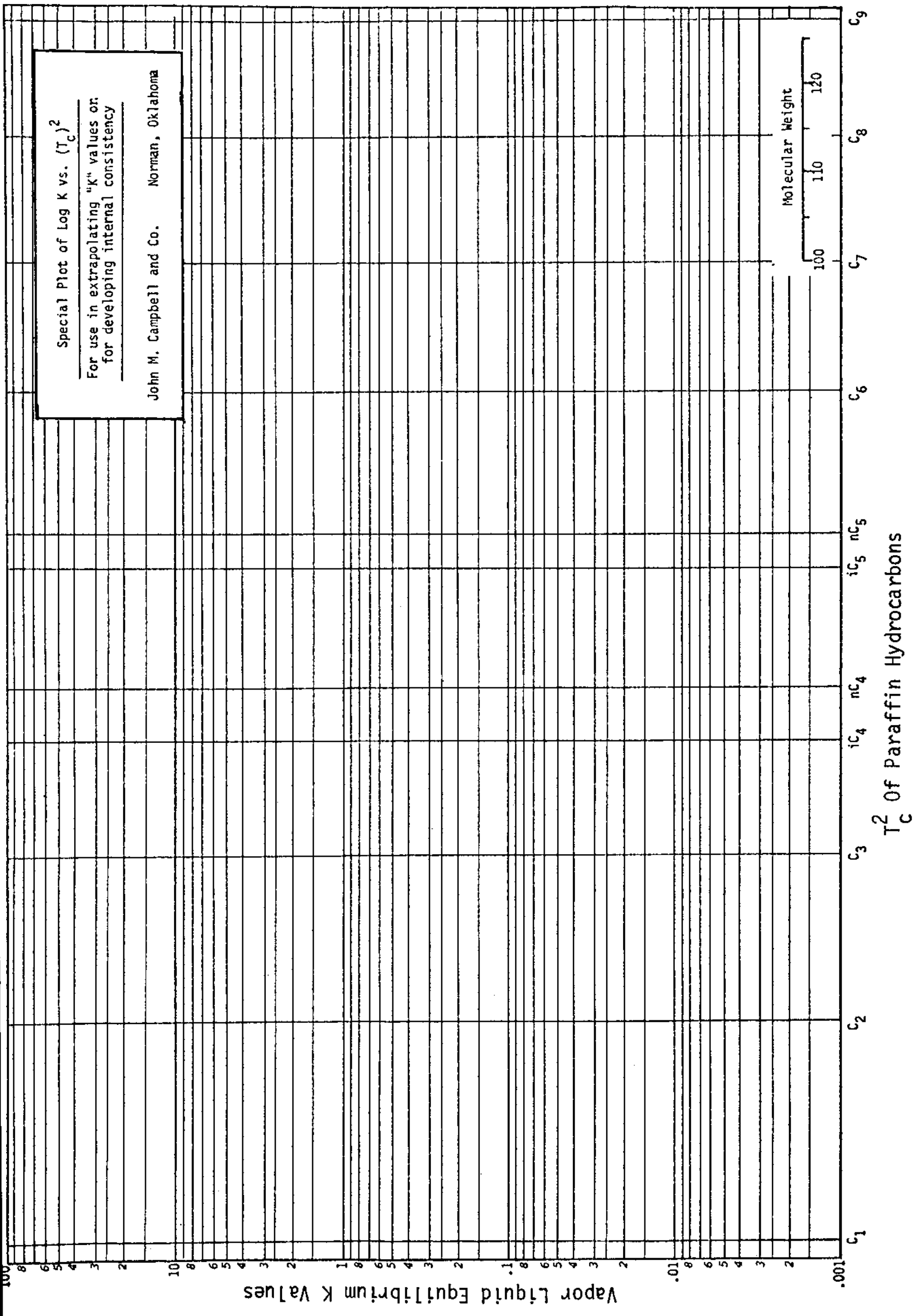


Figure 5.3 Plot of Log K vs. T_c^2

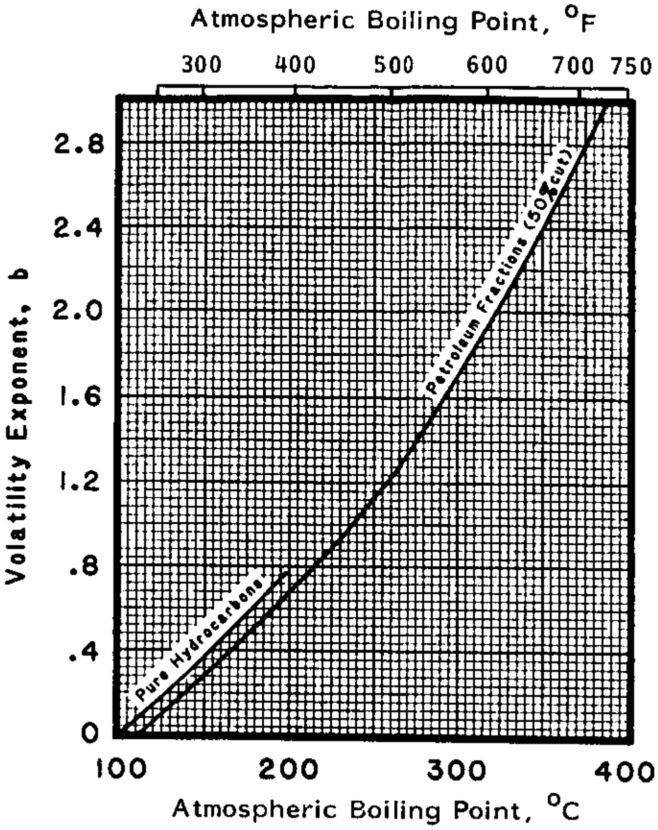


Figure 5.4 Volatility Exponent for Use in Finding the K of C7+

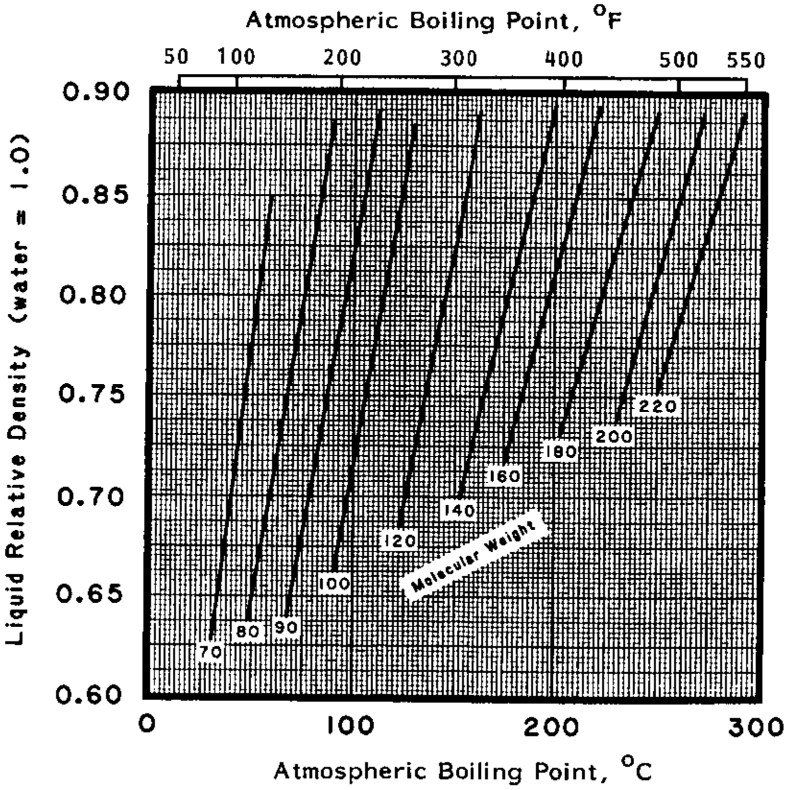


Figure 5.5 Another Correlation for Estimating Hydrocarbon Boiling Point

3. **Use of a KP Plot**

   This is a modification of Method 4. One plots log (KP) vs. b[(1/T_b) – (1/T)] for the components for which K values are known. Extrapolate this straight line to obtain K for the fraction having a boiling point (T_b) in absolute units. This boiling point for a given fraction may be estimated from Figure 5.5. The value of "b" is found from the equation

$$b = \frac{A\,[\log P_c - \log B]}{(1/T_b) - (1/T_c)} \tag{5.8}$$

Where: B = 100 kPa [or 14.50 psia]
A = 1.8 when T is in K and B in kPa [1.0 when T is in °R and B in psia]

4. **Plot log K vs (1/T_b)**

   This is a straight line for the heavier components. It has been superseded for the most part by the first three methods. It also is extrapolated to the T_b of the heaviest fraction.

5. **Use Octane K Value**

   If the heaviest fraction is formed by condensation of liquid from a previously all vapor stream, one usually may approximate the heptanes plus fraction by using the K of octane (or the K of heptane for hexanes plus, etc.) or, as an alternative, using the component K value closest in molecular weight to that of the heaviest fraction.

APPLICATION OF K VALUES

Three different calculations are required for vapor-liquid phase behavior calculations: (1) a bubblepoint calculation to define that portion of the phase envelope, (2) a dewpoint calculation for the remaining portion, and (3) a flash calculation at any P and T inside the phase envelope.

In the discussion which follows, the K values from Appendix 5A will be used in all examples.

Bubblepoint Determination

The bubblepoint may be defined as that condition at which the system is all liquid with one (infinitesimally small) bubble of vapor present. The amount of vapor is specified as a matter of convenience so that the composition of the liquid is essentially equal to the composition of the total system. By definition, the sum of the mol fraction of all components must add up to unity in both vapor and liquid. It follows that

$$\sum K_i\,x_i = \sum y_i = 1.0 \tag{5.9}$$

Where: $K_i x_i$ = K value times "x" the mol fraction for each component in the mixture

The procedure is as follows for a system of known composition:

1. Assume a temperature for the known pressure (or assume pressure if temperature is known).
2. Find K_i at pressure and temperature known and assumed.
3. Multiply K_i from Step 2 by corresponding x (Eqn. 5.9).
4. If summation of values from Step 3 is 1.0, you assumed the right pressure or temperature. If not, repeat Steps 1-3 until $\Sigma K_i x_i = 1.0$, within accuracy limits prescribed.

Note: One can use mol percent instead of mol fraction for x_i, in which case Equation 5.9 sums up to 100 instead of unity.

Example 5.1: etermine the bubblepoint of the mixture shown at 4.0 MPa [580 psia] assuming that the C_7+ fraction has the properties of C_8. This is a trial-and-error solution. One must guess a temperature, find K_i, and calculate $\Sigma K_i x_i$. The right temperature has been assumed when $\Sigma K x_i = 1.0$ (or 100).

Comp.	x_i	K_i	$K_i x_i = y_i$
CO_2	1.25	1.50	1.88
H_2S	0.50	0.41	0.21
C_1	21.36	3.48	74.33
C_2	36.78	0.58	21.33
C_3	10.21	0.16	1.63
iC_4	6.38	0.063	0.40
nC_4	9.84	0.043	0.42
iC_5	2.63	0.017	0.04
nC_5	4.01	0.012	0.048
C_6	3.90	0.0036	0.014
C_{7+}	3.14	0.00033	0.001
			100.30

The solution above is the final try at T = −14°C [7°F]. Since percentages rather than mol fractions are used, the right-hand column should sum to 100.

Dewpoint Determination

The dewpoint is defined as an all vapor system except for one small droplet of liquid. In this case, the system composition, z_i, is equal to that of the vapor, y_i. So,

$$\sum (y_i/K_i) = \sum x_i = 1.0 \tag{5.10}$$

Example 5.2: Determine the dewpoint of the mixture shown at 4.0 MPa [580 psia], assuming the C_{7+} fraction has the properties of C_8.

Comp.	y_i	K_i	$y_i/K_i = x_i$
CO_2	1.25	4.83	0.26
H_2S	0.50	2.38	0.21
C_1	21.36	6.36	3.36
C_2	36.78	2.84	12.95
C_3	10.21	1.57	6.50
iC_4	6.38	1.02	6.25
nC_4	9.84	0.86	11.44
iC_5	2.63	0.55	4.78
nC_5	4.01	0.49	8.18
C_6	3.90	0.28	13.93
C_{7+}	3.14	0.097	32.37
			100.23

The results above are for an assumed temperature of 153°C. It is obtained after a series of trials. Previous trials were made at: 120, 139 and 150°C.

The system in question is a rich one. The percentage methane is low and there are a lot of heavy ends. Notice that in spite of this, methane is the biggest number in the bubblepoint K_ix_i column. If the percentage methane had been larger it would have had an even greater effect.

For the dewpoint, the heptanes plus had the largest single effect. This quantity is the least exact both from the viewpoint of analysis results and its properties. A dewpoint calculation is therefore inherently less exact than a bubblepoint calculation, particularly for lean gases containing only a trivial quantity of these heavy ends.

Many hydrocarbon dewpoint control plants are designed that use some type of refrigeration process. Theoretically, the gas leaving the gas-liquid separator is at its dewpoint (as with any equilibrium separation). In practice, this separator will be designed to operate 10-11°C [18-20°F] below the desired dewpoint temperature. Why? To account for potential error in the dewpoint calculation and for some degree of liquid entrainment which occurs inevitably, at least periodically. The liquid leaving an equilibrium vessel is at its bubblepoint.

Flash Calculation

The purpose of the flash calculation on a two-phase system is to establish the amounts of vapor and liquid, and the analysis of each. At the beginning of this chapter I showed a schematic view of an equilibrium separation.

Overall balance

$$F = V + L$$

For each component

$$F z_i = V y_i + L x_i \tag{5.11}$$

Where:
F = mols of total feed entering separation process
V = mols of gas leaving system for F mols of feed
L = mols of liquid leaving system for F mols of feed
z_i = mols of component "i" in feed stream per mol total feed
y_i = mol fraction of component "i" in gas stream (V)
x_i = mol fraction of component "i" in liquid stream (L)

For convenience, let F = 1 mol, then $z_i = y_iV + x_iL$, but by definition of the equilibrium constant, $y_i = Kx_i$. Substituting accordingly and rearranging the terms,

$$x_i = \frac{z_i}{L + V K_i} \quad , \quad y_i = \frac{z_i}{V + (L/K_i)}$$

Furthermore, the sum of all the mol fractions in any stream must equal 1. Therefore, algebraically,

$$\sum x_i = \sum \frac{z_i}{L + V K_i} = 1.0 \quad , \quad \sum y_i = \sum \frac{z_i}{V + (L/K_i)} = 1.0 \tag{5.12}$$

Therefore, a means of solution is offered, though it is trial and error. By assuming an L or V, one may solve iteratively the equations above. If Σx_i or $\Sigma y_i = 1.0$, a solution is obtained. Thus, the compositions of the streams as well as the quantities may be calculated.

Convergence is obtained in Equation 5.12 when Σx_i or $\Sigma y_i = 1.0$.

Another objective function is $(\Sigma y_i - \Sigma x_i) = 0$. This results in Equation 5.13.

$$\sum y_i - \sum x_i = \sum \frac{z_i (K_i - 1)}{V (K_i - 1) + 1} = 0 \tag{5.13}$$

Equation 5.13 is the most common objective function used in computer programs.

The basic calculation procedure for a flash is as follows:

1. Find K at the pressure and temperature of the two-phase system (separation conditions).
2. Assume V or L.
3. Solve flash equation (Equation 5.12 or 5.13, or equivalent).
4. If objective function is satisfied, calculation is complete. If not, repeat Steps 2-4.
5. Calculate x_i and y_i

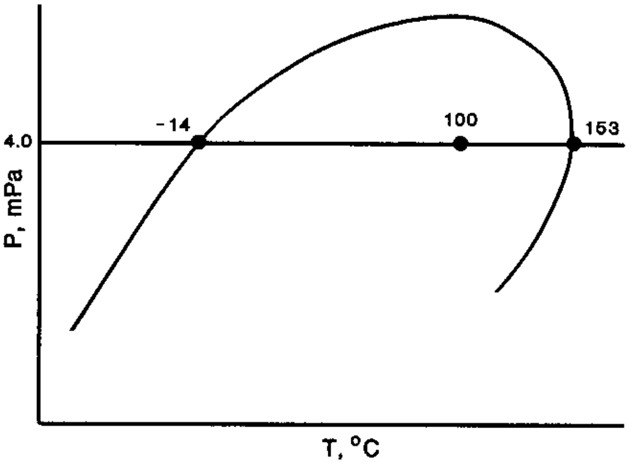
Example 5.3 illustrates the calculation details.

Example 5.3: Calculate the analyses and quantities of vapor and liquid for a flash at 4.0 MPa [580 psia] and 100°C [212°F]. Use the K values in Appendix 5A.

Comp.	z_i	K_i	$y_i - x_i$	x_i	y_i
CO_2	1.25	4.050	1.139	0.373	1.512
H_2S	0.50	1.740	0.236	0.319	0.554
C_1	21.36	5.950	21.974	4.439	26.413
C_2	36.78	2.160	22.535	19.427	41.962
C_3	10.21	1.030	0.299	9.979	10.279
iC_4	6.38	0.600	−3.688	9.220	5.532
nC_4	9.84	0.490	−8.264	16.203	7.940
iC_5	2.63	0.280	−4.250	5.903	1.653
nC_5	4.01	0.240	−7.348	9.668	2.320
C_6	3.90	0.120	−10.646	12.098	1.452
C_7+	3.14	0.031	−11.987	12.370	0.383
		Σ	0.000	100.000	100.000

The 1st step is to assume a V or L. The table reflects the end result of several trials for V = 0.770.

Computer Solutions. These solutions use the same material balance principles shown above. A wide variety of flash equations and convergence algorithms have been used. However, all are not reliable under all phase conditions and constraints imposed by the user. Computer-run dewpoint solutions have sometimes been inaccurate. Volume 3 of this book discusses the computer problem in detail.$^{(5.15)}$



From the flash we have predicted the relative quantity and analysis of the system at the conditions specified. The vapor and liquid leaving are saturated – at their dewpoint and bubblepoint, respectively.

As we can see from the figure, the flash at 100°C is much closer to the dewpoint than the bubblepoint. Therefore, we would expect far more gas than liquid. If we remember the quality lines from Chapter 4, lowering the temperature would lower the relative quantity of vapor.

Determination of Phase

It often is necessary to determine whether the fluid in a line is all gas, all liquid or two-phase. If only this information is needed it is not necessary to make the above detailed calculations. One can set up a calculation for $\Sigma K_i z_i$ and $\Sigma (z_i/K_i)$ at the pressure and temperature of the system.

If: $\Sigma K_i z_i$ and $\Sigma (z_i/K_i)$ are both greater than 1.0 (or 100), a two-phase system exists.

$\Sigma K_i z_i$ is less than 1.0 (or 100), the system is all liquid.

$\Sigma (z_i/K_i)$ is less than 1.0 (or 100), the system is all vapor.

(Both $\Sigma K_i z_i$ and $\Sigma (z_i/K_i)$ cannot be less than 1.0).

If you do this manually, the following column headings are used: Component, z_i, K_i (at P and T), $K_i z_i$ and z_i/K_i. This not a trial-and-error solution.

Stage Separation

Several stages of separation are often used as shown below. This is called either two-stage or three-stage separation, depending on whether you call the stock tank a stage. Most systems consist of two-separator stages, but three or more may be used in selected instances.

The first-stage pressure is normally fixed by specification or economics. The stock tank pressure will be at, or near, atmospheric pressure. Intermediate stage pressures can be fixed arbitrarily to yield maximum stock tank recovery.

One can optimize these intermediate pressures by a series of flashes. The flash in Stage One fixes the analysis and quantity of liquid leaving, which is the feed to Stage Two. A corresponding flash on Stage Two gives stock tank feed.

A flash is normally performed on the stock tank even though the actual process is some combination of differential vaporization, weathering, "breathing" losses and flash vaporization. The result is good enough for planning purposes and the calculation is routine.

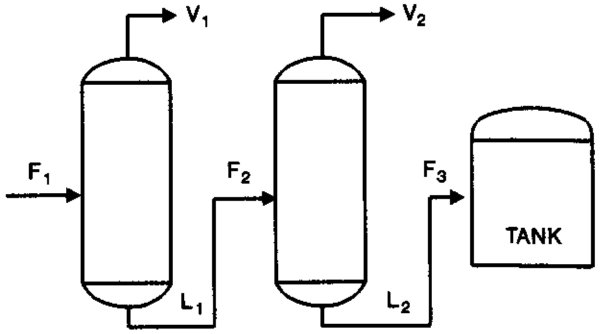
For a separation involving two separators, the optimum second stage pressure may be estimated by the Equations 5.14-5.15.

Pseudo-Relative Density of Feed > 1.0

$$P_2 = \alpha A (P_1)^{0.686} + C \qquad (5.14)$$

Pseudo-Relative Density of Feed < 1.0

$$P_2 = \alpha' A (P_1)^{0.765} + C' \qquad (5.15)$$



Where:		Metric	English
P_1	= 1st stage pressure	kPa	psia
P_2	= 2nd stage pressure	kPa	psia
α	= constant	1.83	1.0
α'	= constant	1.57	1.0
C	= shifting constant	(A+0.057)/0.0034	(A+0.057)/0.0233
C′	= shifting constant	(A+0.028)/0.0017	(A+0.028)/0.012
A	= constant found from Fig. 5.6		

Example 5.4: Find the optimum second-stage pressure for the wellstream shown below for P_1 = 3.45 MPa [500 psia].

(1) Comp.	(2) mol fraction	(3) mol weight	(4) (2)x(3)
C_1	0.40	16.01	6.40
C_2	0.20	30.07	6.01
C_3	0.10	44.09	4.41
C_4	0.10	58.12	5.81
C_5	0.10	72.15	7.22
C_6	0.05	86.17	4.31
C_{7}+ (C_8)	0.05	115.22	5.76
	1.00		39.92

Relative Density = 39.92/28.96 = 1.38
$C_1 + C_2 + C_3$ = 70 percent

From Fig. 5.6, A = 0.42; hence, from Eq. 5.14

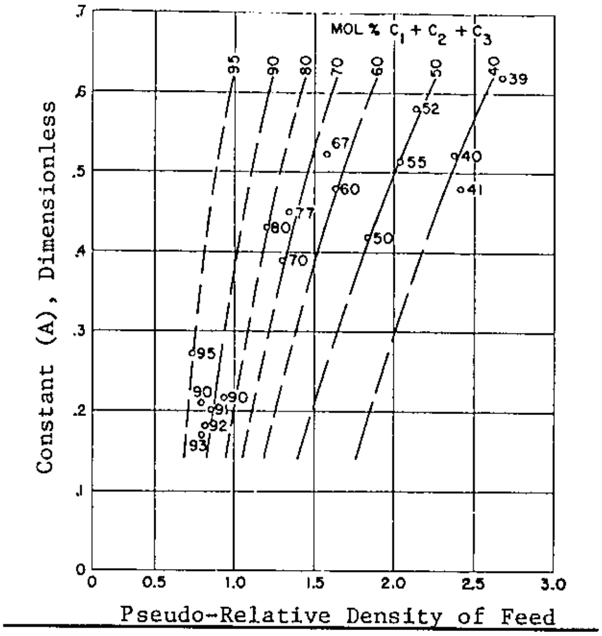


Figure 5.6 Constant "A" vs. Pseudo-Density of Feed at T=80°F

Metric: $C = \dfrac{(0.42 + 0.057)}{0.0034} = 140$ $P_2 = (1.83)(0.42)(3450)^{0.686} + 140 = \underline{\underline{345 \text{ kPa}}}$

English: $C = \dfrac{(0.42 + 0.057)}{0.0233} = 20.5$ $P_2 = (1.0)(0.42)(500)^{0.686} + 20.5 = \underline{\underline{50.4 \text{ psia}}}$

This approach is for estimation purposes only. The most rigorous approach is to optimize pressure using a series of stage separation calculations on a computer.

SOME CONVERSION CALCULATIONS

In making equilibrium calculations it often is necessary to convert or combine quantities to meet the need of that calculation. Summarized below are some convenient calculations often encountered.

Table 5.1 summarizes useful conversion factors for the paraffins through normal decane. In this table standard conditions are:

Metric: 15°C and 100 kPa **English:** 60°F and 14.7 psia

In the preparation of the quantities shown at these standard conditions, the following conversion factors are used:

379.3 scf/lb mol, 23.96 m^3/kmol

Density of water = 8.33 lbm/U.S. gal = 1000 kg/m^3

The values shown in Table 5.1 can be calculated for any hydrocarbon fraction if one knows the molecular weight and specific gravity of that fraction. Small deviations in some numbers result from round-off procedures.

TABLE 5.1
Properties of Paraffin Hydrocarbon Liquids - Metric Units

Component	Mol Wt.	std m^3 Gas per m^3 Liquid	m3/kmol	kmol/m^3	Normal BP °C	Relative Density Liquid
Methane	16	(442)	(0.05)	(20)	–162	(0.3)
Ethane	30	281	0.084	11.90	–89	0.356
Propane	44	272	0.087	11.49	–42	0.508
i-Butane	58	229	0.103	9.71	–12	0.563
n-Butane	58	238	0.099	10.10	–0.5	0.584
i-Pentane	72	205	0.116	8.62	28	0.624
n-Pentane	72	207	0.114	8.77	36	0.631
n-Hexane	86	182	0.130	7.69	69	0.664
n-Heptane	100	162	0.146	6.85	98	0.688
n-Octane	114	146	0.162	6.17	126	0.707
n-Nonane	128	133	0.177	5.64	151	0.722
n-Decane	142	122	0.193	5.17	174	0.734

Note: std m^3 @ 15°C, 100 kPa

TABLE 5.1(a)
Properties of Paraffin Hydrocarbon Liquids - English Units

Component	Mol Wt.	scf/U.S. gal	U.S. gal per lb-mol	lb-mol per U.S. gal	Normal BP °F	Relative Density Liquid
Methane	16	(59.1)	(6.40)	(0.1555)	–259	(0.3)
Ethane	30	37.5	10.12	0.0988	–127	0.356
Propane	44	36.4	10.42	0.0959	–44	0.508
i-Butane	58	30.7	12.38	0.0808	11	0.563
n-Butane	58	31.8	11.94	0.0836	31	0.584
i-Pentane	72	27.4	13.86	0.0722	82	0.624
n-Pentane	72	27.7	13.71	0.0729	97	0.631
n-Hexane	86	24.4	15.59	0.0641	156	0.664
n-Heptane	100	21.7	17.46	0.0573	209	0.688
n-Octane	114	19.6	19.38	0.0516	258	0.707
n-Nonane	128	17.8	21.28	0.0470	303	0.722
n-Decane	142	16.3	23.20	0.0431	345	0.734

Composition Expressed as Liquid, Mol or Weight %

Analyses are given normally in mol percent, but it is sometimes necessary to convert from one type to another. The procedure is illustrated below.

Comp.	(1)	(2)	(3)	(4)	(5)	(6)	(7)
	Mol %	Mol Wt	(3) = (1)x(2)	Wt %	γ_L	(6) = (3)/(5)	Liquid %
C_1	5.40	16	86.4	0.74	0.30	288	1.79
C_2	6.98	30	209.4	1.80	0.36	582	3.61
C_3	12.54	44	551.8	4.74	0.51	1082	6.72
iC_4	5.38	58	312.0	2.68	0.56	557	3.46
nC_4	6.42	58	372.4	3.20	0.58	642	3.98
C_5+*	63.18	160	10 108.8	86.84	0.78	12 960	80.44
	100.00		11 640.8	100.00		16 111	100.00

* MW = 160, γ_L = 0.775

Column (3) is the mass (in kg or lbm) corresponding to the mols in Column (1). Column (4) is simply each entry in Column (3) divided by the total of that column. Column (6) is each mass entry in Column (3) divided by relative density to find the equivalent volume. Each entry in Column (6) divided by its total gives the values shown in Column (7).

For a gas,

$$\text{mol \%} = \text{volume \%} = \text{partial pressure \%}$$

Composite Stream Analysis

The best way to sample a two-phase stream is to use a test separator. In this you sample and analyze the liquid and gas streams leaving, as well as measure their flow rates. These results may then be recombined to form a composite (or total) stream analysis.

To do this the flow rates must be expressed in mass units before they can be added. Since no chemical reaction is occurring, the mol is a convenient mass unit. The example following illustrates the procedure.

Example 5.5:

Gas flow rate: 67 800 std m^3/d Pressure: 6.9 MPa
Temperature: 46°C
Gas and liquid analysis: As shown below
Gas oil ratio: 2335 m^3 gas/m^3 oil
Mol wt separator liquid: 80.6 (from analysis); separator liquid, γ_L = 0.7

Metric Unit Solution: **(0.7) x 1000 kg/m^3** = 700 kg/m^3 oil (oil density)
700/80.6 = 8.685 kmol oil/m^3
Gas-oil ratio (from test) = 2335 m^3 gas/m^3 oil or 428 m^3 oil/10^6 std m^3
8.685 kmol/m^3 x 428 $m^3/10^6m^3$ = 3717 kmol oil/10^6 std m^3

Note: 10^6 std m^3 = 41 740 kmol of gas

Example 5.5 (Cont'd.):

In tabular form:

(1) Comp.	(2) Mol fr. Gas	(3) $\frac{\text{kmol}}{10^6 \text{ std m}^3}$	(4) Mol fr. Liq.	(5) $\frac{\text{kmol}}{10^6 \text{ std m}^3}$	(6) Well Stream kmol	(7) Mol Fr. Well Stream
C_1	0.8946	37 341	0.2022	752	38 093	0.8380
C_2	0.0541	2 258	0.0656	244	2 502	0.0550
C_3	0.0310	1 294	0.1116	415	1 709	0.0376
iC_4	0.0055	229	0.0445	165	394	0.0087
nC_4	0.0078	326	0.0708	263	589	0.0130
iC_5	0.0022	92	0.0465	173	265	0.0058
nC_5	0.0012	50	0.0337	125	175	0.0038
C_6	0.0026	109	0.1033	384	493	0.0108
C_7	0.0010	42	0.3218	1196	1 238	0.0272
	1.0000		1.0000		45 458	1.0000

The above metric tabulation and the corresponding English unit tabulation are based on the same principle. Columns (2) and (4) are the gas and liquid analyses from sampling. Multiplying each number in Column (2) by 41 740 gives the number of mols of each component per million std cubic meters of gas shown in Column (3). Multiplying each number in Column (4) by 3717 kmol oil/million std cubic meters gives the numbers shown in Column (5).

Cols. (3) and (5) may now be added (for they are on the same basis) to obtain Col. (6). Dividing each entry in Col. (6) by its total gives the desired mol fraction analysis of the well stream.

Conversion of Flash Results for Actual Flow Rates

It is customary to make flash calculations on the basis of 1.0 or 100 mols of feed because it is convenient. Once V and L have been obtained on this basis, they must be adjusted to find the actual rate.

If liquid is obtained from a separation, that liquid existed in the feed because none is formed in the separator. The amount of feed will be the quantity of liquid plus the quantity of gas produced. Sometimes this feed flow rate is expressed in gas volumetric units even though some liquid is present. If so, remember that this is a *pseudo* or *equivalent volume* and not the real volume of the two-phase stream. In reality, one normally should express the flow rate in a convenient mass unit to avoid the volume problem.

The traditional gas flow terminology originated in the U.S. The basic unit was the standard cubic foot; the volume of gas is measured at 60°F and a standard pressure which has varied from 14.4 psia to 15.025 psia in various situations. For process calculations it has been common to use 14.7 psia as the base or standard pressure.

In metric countries 0°C and 15°C have been used commonly as base temperatures. The base pressure has been expressed in atmospheres, mm or cm of mercury, kg/cm^2 or bar. With the advent of the SI System, 100 kPa (1 bar) is the logical base pressure and will be used throughout this book. Some have proposed using 101.325 kPa (1 atm) as a base, but it is an inconsistent ("bastard") number in a "powers of ten" system.

In this book we will use the standard cubic meter as the basic gas unit, with the base conditions being 100 kPa and 15°C. The standard SI prefixes will be used. So, one million cubic meters will be written as

10^6 or (E+06). To differentiate *standard* from *actual* cubic meters, the standard volume will be written as "std m^3." The subscript "sc" also will be used to denote standard conditions.

In English units a standard cubic foot will be measured at 60°F and 14.7 psia in the traditional manner. One million standard cubic feet will be written as 1 MMscf/d in this book. In some references you will find it written as MMSFD, MMscfD, MMcfD, etc. These all mean the same thing.

> *Note:* The prefix M (mega) is one million in the metric system. In the English system it is the Roman numeral for 1000. This is a possible source of confusion. In this book, to prevent confusion, I *will not* use the metric prefix mega (M) to describe gas volumes.

The liquid volumes will be expressed in cubic meters whenever practical. For small volumes the liter will be used. It will be abbreviated L in this text.

In English units the term gallon (U.S. or Imperial) will be used seldom. The cubic foot is preferred. In some cases the API barrel (bbl) will be used because it is employed commonly around the world. It will, however, be related to cubic meters or cubic feet whenever practical.

Basic Conversion Factors

Metric Units

100 000 N/m^2 = 100 000 Pa = 1.02 kg/cm^2 = 100 kPa

Standard conditions: 15°C and 100 kPa

1 000 000 std m^3 = 10^6 std m^3 = 41 740 kmol

There are 23.96 std m^3/kmol and 41 740 kmol/10^6 std m^3

Density of air at 15°C and 100 kPa = 1.21 kg/m^3

Density of gas at 15°C and 100 kPa = $(\gamma_g)(1.21)$

1 m^3 = 1000 liters; 1 liter = 1000 cm^3

1 g/cm^3 = 1000 kg/m^3

English Units

Standard conditions: 60°F and 14.7 psia

1 000 000 std cu ft = 1 MMscf = 2636 lb mol

There are 379 std ft^3/lb mol at 60°F and 14.7 psia

Density of air at 60°F and 14.7 psia = 0.0764 lb/ft^3

1 ft^3 = 7.48 U.S. gal = 6.23 Imperial gal

1 bbl = 42 U.S. gal = 35 Imperial gal = 5.61 ft^3

Conversions between Units

1 std ft^3 (@ 60°F and 14.7 psia) = 0.0286 std m^3 (@ 15°C and 100 kPa)

1 std m^3 (@ 15°C and 100 kPa) = 34.92 std ft^3 (@ 14.7 psia and 60°F)

1 bbl = 159 L = 0.159 m^3

1 m^3 = 35.31 ft^3 = 264 U.S. gal = 200 Imperial gal = 6.29 bbl

1 L = 0.264 U.S. gal = 0.22 Imperial gal = 0.0353 ft^3

ACCURACY OF EQUILIBRIUM CALCULATIONS

The accuracy of the results of calculations involving K values depends on the reliability of the sampling, of the analysis of that sample, and on the K value correlation used. The importance of sampling and analysis have been discussed previously.

There is no single K value correlation that is superior for all mixtures encountered. A knowledgeable practitioner may have two or three different models or programs available. Generally, crude oil and NGL phase behavior is handled by different models.

All K values are sensitive to composition, particularly the very volatile components like nitrogen, methane and ethane. In correlations like those shown in Appendixes 5A and 5B, the potential error in K for nitrogen and methane may be very large. Notice how different the values are for methane (at the same conditions) in these two Appendixes.

For design purposes, several models may be used to determine a range of results. This range, rather than one set of "magic" numbers, is then used to specify equipment. The name of the game is flexibility. It is doubtful if one ever will encounter the analyses, flow rates and exact other conditions specified.

It is most important that the K values be internally consistent. One can use Figure 5.3 for this purpose or accomplish it mathematically. Establish the best straight line through the points for propane through hexanes. Extrapolate this line to find the K values for methane and ethane and for the heavier ends. Maintaining the proper relativity of the K values reduces the effective error of the results.

An experienced person usually can predict the *quantity* of a specified liquid within ±5-6% (for a specified analysis and conditions). The detailed analysis will be off more than this. This is important, for in many systems a series of VLE calculations is made; the output from one is the input to another. The errors thus cumulate. Many less than desirable systems result from failure to recognize this.

Table 5.2 shows the comparative results of a low temperature calculation by several K value programs. Other comparisons are shown in the literature.(5.17) Comparisons like these should be made for your specific systems.

PRODUCT SPECIFICATION

In a production system using stage separation, the final liquid product is fixed by the stock tank conditions. Although the VLE mechanism in this tank is more nearly differential vaporization rather than flash vaporization, the latter is most often used for calculation purposes.

Many liquid products have a vapor pressure specification which must be satisfied by separation or stabilization.

Reid Vapor Pressure

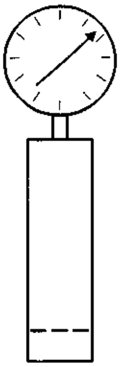
In order to measure true vapor pressure there can be nothing in the test cell except the liquid being tested. In practice this means that any air present at the beginning of the test must be removed (the cell evacuated) before introduction of the sample. This is inconvenient in a field or plant laboratory using operating personnel.

Reid vapor pressure (RVP) is the substitute specification used for some NGL liquids. If you see RVP after the vapor pressure it is the Reid vapor pressure; if RVP is not shown it is a true vapor pressure, by default.

TABLE 5.2
Comparison of Experimental and Predicted Liquefaction

	Weight % Total Fluid Condensed	Weight % Total Ethane Condensed	Weight % Total Propane Condensed
–55°F & 1000 psia			
Experimental	17.12	25.6	45.3
BWRS	18.16	28.0	47.4
Peng-Robinson	18.67	29.2	51.1
GPA SRK	19.01	29.9	52.04
–60°F & 900 psia			
Experimental	19.02	30.9	56.4
BWRS	19.57	32.3	56.5
Peng-Robinson	19.66	32.8	58.9
GPA SRK	19.83	33.3	60.2
–70°F & 1000 psia			
Experimental	33.8	44.6	57.1
BWRS	32.5	44.1	57.7
Peng-Robinson	31.0	44.7	62.2
GPA SRK	31.2	45.1	63.0
–100°F & 500 psia			
Experimental	23.23	52.1	85.4
BWRS	24.7	55.5	86.7
Peng-Robinson	24.9	56.6	87.9
GPA SRK	24.5	56.1	88.2

An RVP pressure must be determined in a *standard cell* to properly take care of the air partial pressure. The ratio of vapor space to sample space must be correct so the correction charts between Reid and true vapor pressure are applicable.



If RVP is known it may be corrected to an equivalent true vapor pressure. Figure 5.7 is convenient for this purpose, and accurate enough for most applications. True vapor pressure is read from the left-hand ordinate at the temperature of the sample.

Figure 5.8 is a nomograph for crude oils that has been modified from one issued by the API. It is used commonly for accepting crude oil shipments. Many purchasers have specifications on such a shipment to prevent loss of liquid in storage tanks. The procedure is simple. The temperature and RVP of the oil is measured at the transfer point. Figure 5.8 is then used to determine true vapor pressure to ascertain if that shipment fulfills contract specifications. Tanker imports of crude oil must often have an RVP limit of 10-11 psi.

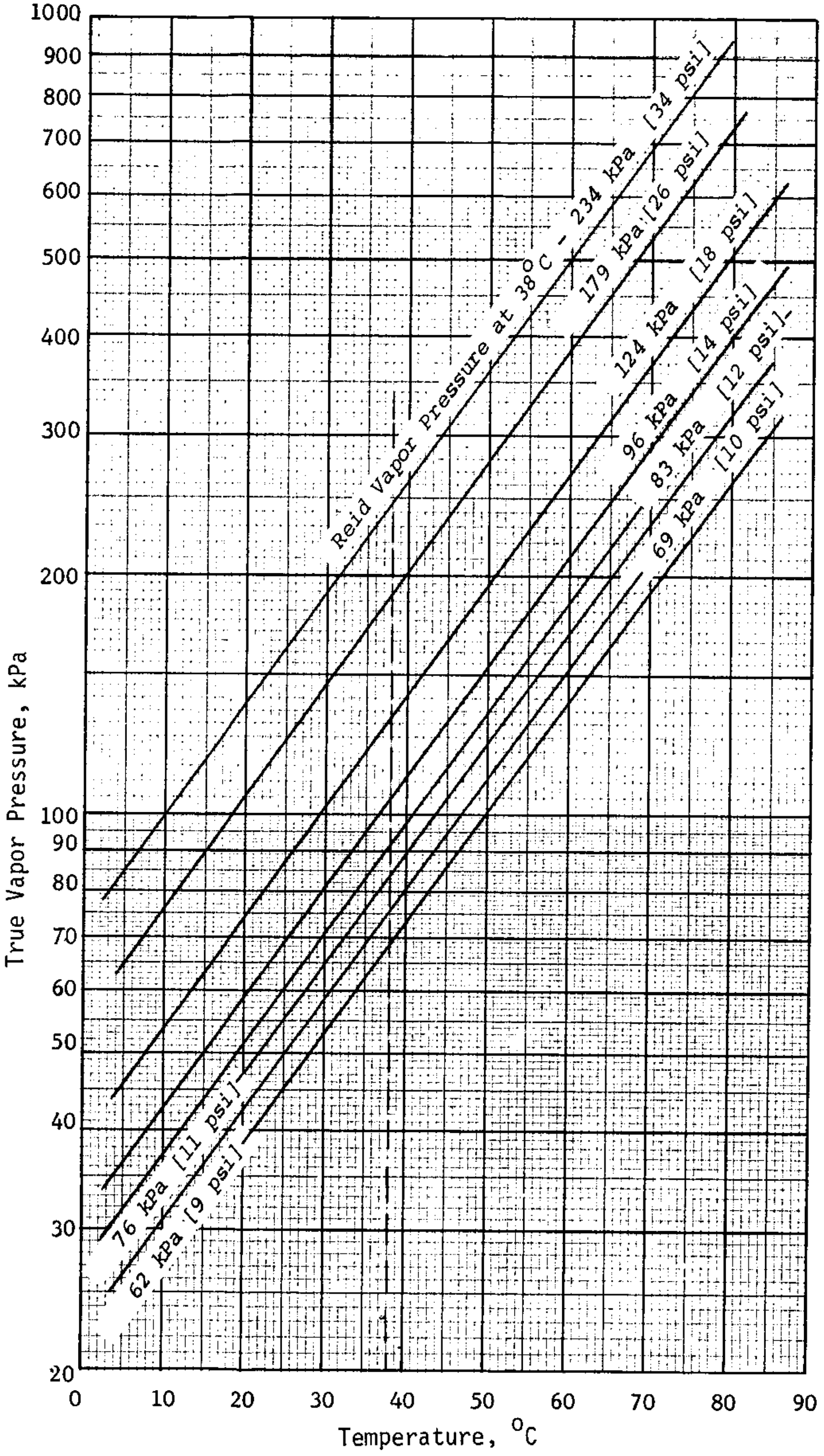


Figure 5.7 Reid Vapor Pressure vs. Temperature for Typical NGL Liquids

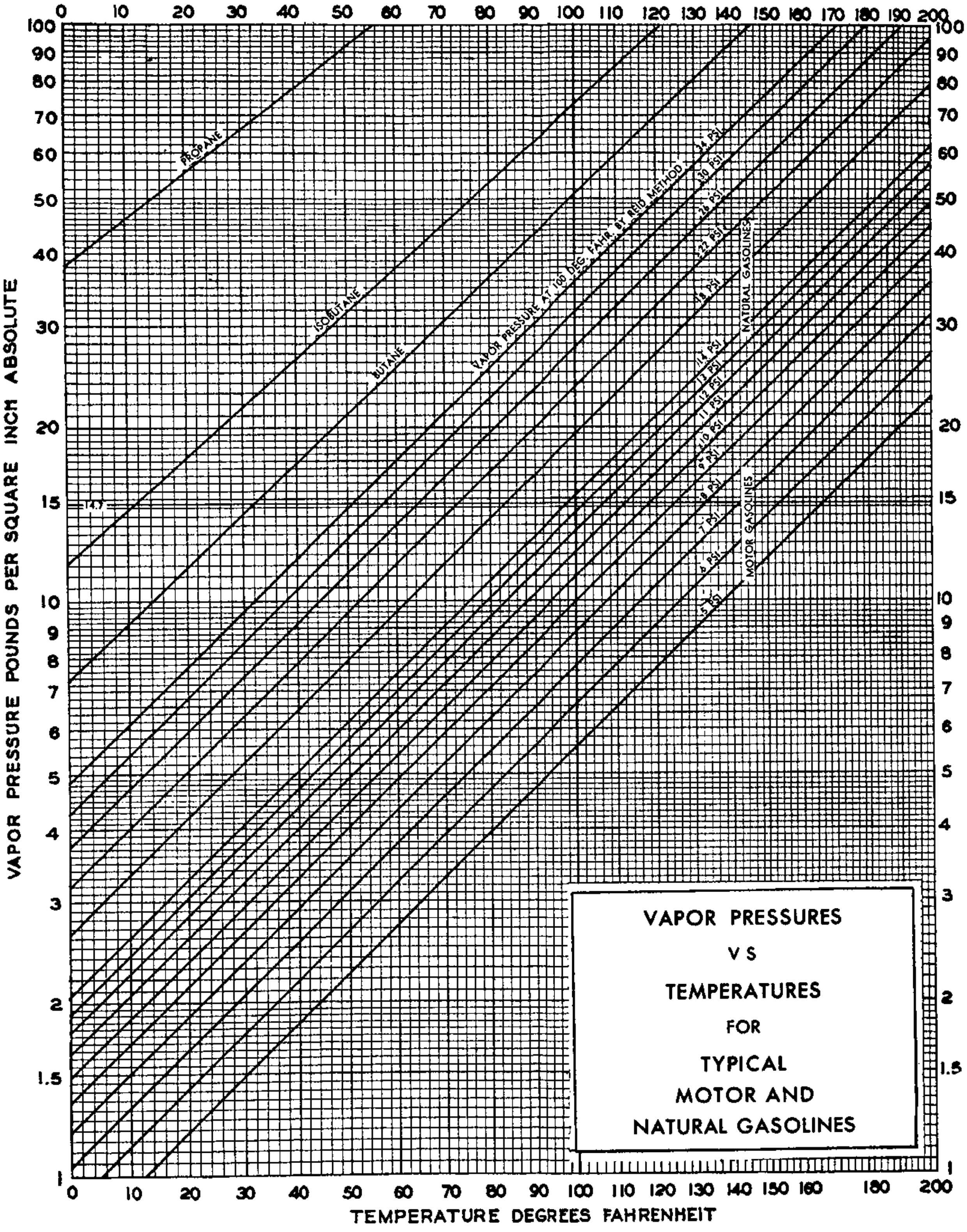


Figure 5.7a Vapor Pressures vs. Temperatures

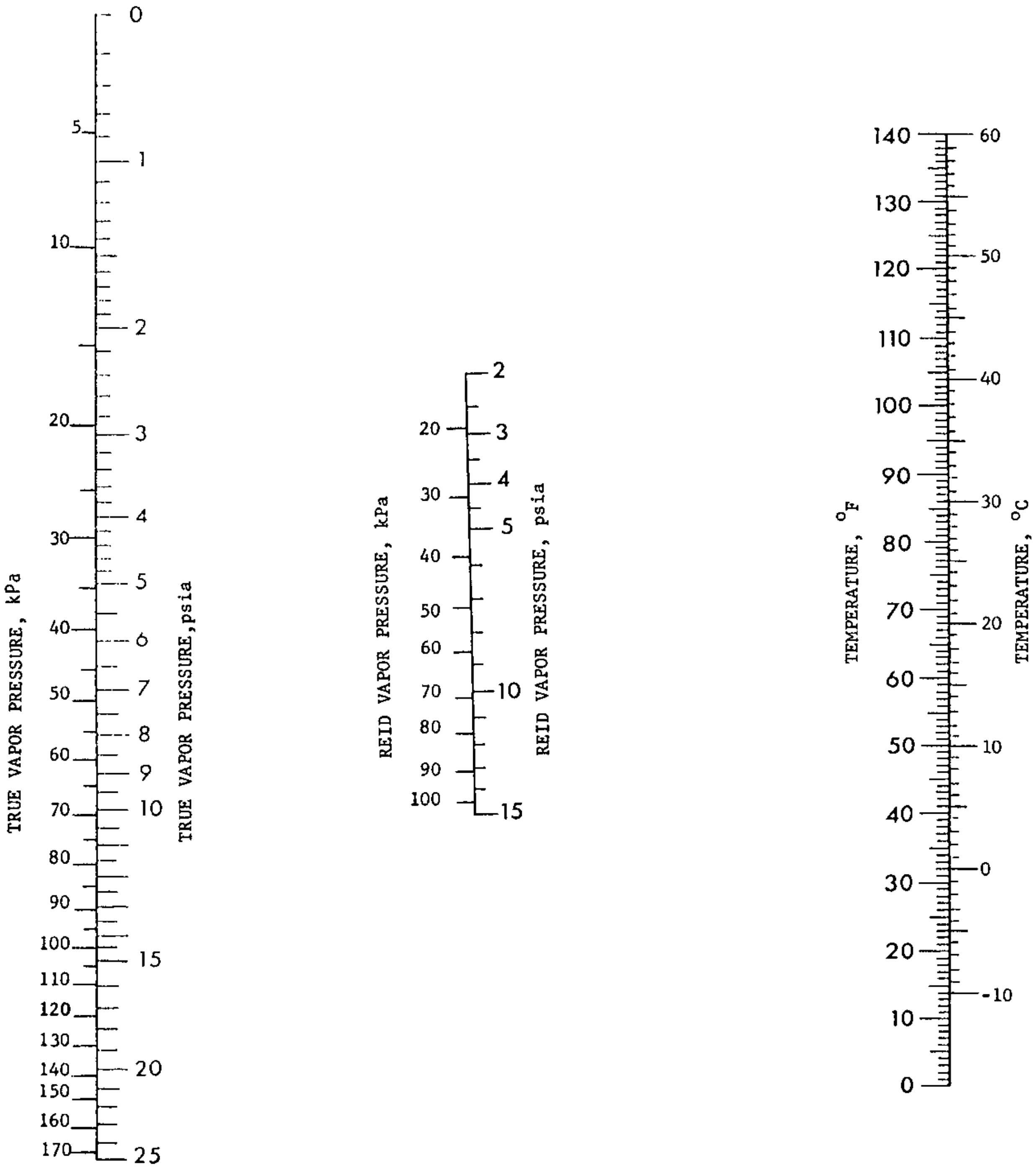


Figure 5.8 Reid Vapor Pressure vs. True Vapor Pressure for Crude Oils

PRELIMINARY PRODUCT SPLITS

Very early in the planning phase, before detailed calculations are made, it is necessary to establish the approximate quantity of the fluids to be produced. If one of the specifications is a vapor pressure the procedure which follows is useful approximation for planning and specification purposes.

The following general procedure is normally employed in making this preliminary split.

> *Natural Gasoline.* It is assumed that all components having a lower vapor pressure than that specified for the product will appear 100% in that product. For almost all natural gasolines this implies that all of the available pentanes plus entering the fractionation system will appear in the final product. One then adds the proper amount of butanes to give the desired vapor pressure. In the unusual case where all of the butanes won't do this, the necessary amount of propane is added.
>
> *Commercial Propane.* Simply calculate the volume of propane available from percent propane in the potential product stream. The actual product will have small percentages of ethane and butane, but these will have little effect on the quantity of product.
>
> *Commercial Butane.* Same as above except based on total availability of iso- and normal butane.

The vapor pressure is the primary property used to make this split. It is assumed that the mol fraction of each component times its vapor pressure represents the contribution of that component to the total mixture vapor pressure. The sum of this multiplication for each component is the total mixture vapor pressure (P_v).

$$P_v\,(\text{mixture}) = \sum x_i P_{vi} \tag{5.16}$$

The normal fractionator can make a reasonably sharp separation between hydrocarbons with different molecular weight – butane from pentane, pentane from hexane, etc. However, a sharp separation between isomers possessing the same molecular weight is difficult – i-butane from n-butane and i-pentane from n-pentane. What separation that occurs will depend on the not yet available detailed design of the fractionator.

In the making of this preliminary split, Equation 5.17 will prove useful for predicting isomer behavior.

$$\frac{L'}{L} = A\left(\frac{x'}{x}\right) \tag{5.17}$$

Where:
L = mol of the iso-paraffin in the final mixture
L' = mol of the normal paraffin in the final mixture
x' = mol fraction of normal paraffin in fractionator feed
x = mol fraction of iso-paraffin in fractionator feed
A = 1.25 for butanes and 1.12 for pentanes

The calculation is not truly trial-and-error but one does have to make some kind of guess about what the lightest component will be in desired mixture.

The amount of product produced can be found from the absolute feed rate. The above calculation predicts 61.94 mol of 124 kPa [18 psia RVP] product per 100 mol of feed. This ratio permits conversion of feed rate to product rate in the units of your choice.

This preliminary split will not check actual plant performance exactly but will be very adequate for planning purposes. In most cases the error should not exceed about 5%.

Example 5.6: Calculate the analysis of 124 kPa RVP [18 psi RVP] liquid product at 38°C [100°F]. From Figure 5.7 the true vapor pressure (TVP) is 133 kPa [19.5 psia]. The vapor pressures are from Chapter 3. The analysis of the feed is shown in Column 2 below.

(1)	(2)	(3)	(4)	(5)	(6)	(7)
Component	Mol %	Mols for Split	P_{v_i}	(3)x(4)	Mol Product	Mol %
Propane	33.09	–	1309	–		
i-Butane	8.52	L	498	498 L	5.66	9.14
n-Butane	12.70	1.87 L (= L′)	356	666 L	10.59	17.10
i-Pentane	5.63	5.63	141	794	5.63	9.09
n-Pentane	3.73	3.73	108	403	3.73	6.02
Hexane	10.43	10.43	34.2	357	10.43	16.84
Heptanes + (C_8)	25.90	25.90	3.7	96	25.90	41.81
	100.00	2.87 L + 45.69		1164 L + 1650	61.94	100.00

$$L' = 1.25\left(\frac{12.70}{8.52}\right)(L) = 1.87\ L \text{ for use in Column (3)}$$

$$\Sigma\, x_i\, P_{v_i} = \frac{\Sigma\,(\text{Mol Comp.})(P_{v_i})}{\text{Total Mol}} = \frac{\Sigma\,(\text{Mol Comp.})(P_{v_i})}{2.87\ L + 45.69} = \frac{1164\ L + 1650}{2.87\ L + 45.69} = 133\ , \quad L = 5.66$$

Σ (Mol Comp.)(P_{v_i}) is the total of Column (5).

In English units the results in Column (7) would be the same. The only difference would be the effect of using P_{v_i} values in Column (4) in psia which in turn would affect Column (5).

STORAGE OF LIQUIDS

Proper storage is of economic importance. The loss of potentially salable liquid is not large percentagewise but can represent a substantial amount of money on large systems.

LNG is normally stored at, or near, atmospheric pressure as noted in a later chapter. NGL liquids normally require pressure storage in a spherical or cylindrical tank. Liquid sold as crude oil is usually stored very near to atmospheric pressure.

Pressure Storage

The working pressure of the tank must be higher than the true vapor pressure of the liquid being stored. This tank is, in effect, a vapor pressure cell containing liquid and a vapor phase. The TVP must be calculated at the temperature of the liquid surface and in the vapor space, which will vary.

It also is important that the working pressure be high enough to prevent breathing losses. Thus, a further consideration is the pressure setting of the vacuum vent.

The following nomenclature is used in the storage pressure Equations 5.18 and 5.19 which follow:

		Metric	English
Where:	P_{st} = required storage pressure	kPa(g)	psig
	P_{max} = TVP at max. liquid temp.	kPa(a)	psia
	P_{min} = TVP at min. liquid temp.	kPa(a)	psia
	P_v = pressure at which vacuum vent opens	kPa(a)	psia
	T_{max} = max. avg. temp. of vapor	K	°R
	T_{min} = min. avg. temp. of vapor	K	°R
	A = atmospheric pressure	kPa(a)	psia

For the condition where P_{min} is less than P_v,

$$P_{st} = P_{max} + \left[(P_v - P_{min}) \left(\frac{T_{max}}{T_{min}} \right) \right] - A \tag{5.18}$$

For the condition where P_{min} is greater than P_v,

$$P_{st} = P_{max} - A \tag{5.19}$$

With the situation represented by Equation 5.18, air or gas is admitted to the vapor space through the vacuum vent. Introduction of air can be a safety problem if an explosive mixture results. For this reason, a gas vent system may be employed. For the situation represented by Equation 5.19, air may be purged and kept out of the tank.

Crude Oil Storage

Oil may be stored in a fixed-roof tank of constant volume or a floating head (variable volume) tank. The latter is used to minimize breathing losses and those losses which occur by virtue of the filling method. If the fixed-volume tank is filled from the bottom, some stripping of the liquid already there occurs as gas "breaks out" of the entering oil. If the tank is filled at the top, some splashing or agitation may occur to cause excess liquid entrainment.

The cause of breathing is illustrated by Figure 5.9. A type of breathing also occurs when the tank is being emptied. Air or gas must be admitted to keep the tank from collapsing. Some of the oil must vaporize to maintain an equilibrium mixture.

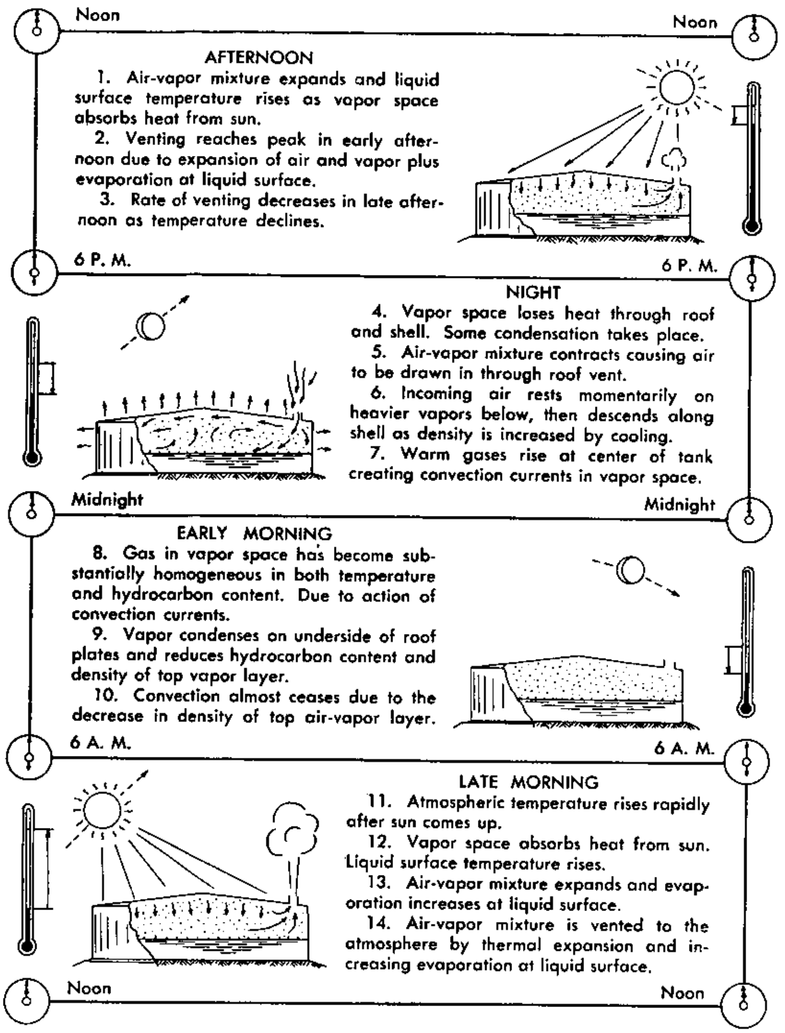
If this loss is too great, some alternative to a simple, fixed volume tank is indicated. One modification uses a layer of small spheres which float on the surface of the oil to form a barrier between the oil and gas. Another alternative is a vapor recovery system. (This usually is a refrigeration system operating on the very rich, effluent tank vapors.) A floating head tank is used for most large storage volumes.

The process of loss involves several mechanisms and thus use of vapor-liquid theory is limited to predict said loss. An API study committee has developed some empirical correlations for predicting oil tank losses from fixed-volume tanks. The key equations from their study are shown below.

The approximate volume loss is estimated from the equation

$$\text{Volume loss in \%} = (m)(\text{Gravity loss in °API}) \tag{5.20}$$

Where:
m = 4.45 for 15° API oil (0.97 g/cm^3)
m = 3.50 for 20° API oil (0.93 g/cm^3)
m = 2.72 for 25° API oil (0.90 g/cm^3)
m = 2.20 for 30° API oil (0.88 g/cm^3)
m = 1.76 for 35° API oil (0.85 g/cm^3)
m = 1.48 for 40° API oil (0.83 g/cm^3)



AFTERNOON

1. Air-vapor mixture expands and liquid surface temperature rises as vapor space absorbs heat from sun.
2. Venting reaches peak in early afternoon due to expansion of air and vapor plus evaporation at liquid surface.
3. Rate of venting decreases in late afternoon as temperature declines.

NIGHT

4. Vapor space loses heat through roof and shell. Some condensation takes place.
5. Air-vapor mixture contracts causing air to be drawn in through roof vent.
6. Incoming air rests momentarily on heavier vapors below, then descends along shell as density is increased by cooling.
7. Warm gases rise at center of tank creating convection currents in vapor space.

EARLY MORNING

8. Gas in vapor space has become substantially homogeneous in both temperature and hydrocarbon content. Due to action of convection currents.
9. Vapor condenses on underside of roof plates and reduces hydrocarbon content and density of top vapor layer.
10. Convection almost ceases due to the decrease in density of top air-vapor layer.

LATE MORNING

11. Atmospheric temperature rises rapidly after sun comes up.
12. Vapor space absorbs heat from sun. Liquid surface temperature rises.
13. Air-vapor mixture expands and evaporation increases at liquid surface.
14. Air-vapor mixture is vented to the atmosphere by thermal expansion and increasing evaporation at liquid surface.

Figure 5.9 Mechanism of Tank Breathing Loss

The actual loss will depend on prior conditioning of the oil, the method and rate of filling and the ratio of liquid surface area to liquid volume. The effects of all of the above losses are summarized in References 5.19 and 5.20. These references contain the same basic information.

The calculation of losses involves many factors, but the two equations which follow are useful approximations of the breathing loss and filling loss for fixed, cone roof tanks.

Breathing Loss. The basic equation is

$$B = \left[\frac{(P)(D)^{1.8}}{A}\right](F_o)(F_p) \qquad (5.21)$$

			Metric	English
Where:	B	= annual breathing loss	m^3	API bbl
	D	= tank diameter	meters	feet
	P	= TVP at average liquid temperature	kPa(g)	psig
	A	= unit factor	74	14.5

F_p = paint factor = 1.0 for aluminum; 0.75 for chalking white; 1.1 for light gray; 1.25 for black; no paint and tank needing repainting

F_o = outage factor based on the average distance to the top flange of the tank found from the following table

Outage		F_o	Outage		F_o
m	ft		m	ft	
0.31	1	0.39	9.15	30	1.23
1.53	5	0.55	10.68	35	1.33
3.05	10	0.72	12.20	40	1.43
4.58	15	0.87	13.73	45	1.53
6.10	20	1.00	15.25	50	1.62
7.63	25	1.12			

Equation 5.21 is based on a tank being about half full on the average, when storing a 65 kPa [9.5 psia] RVP product. Unfortunately, predicting breathing losses may show a 25% variation because of the many factors that cannot be accounted for in a quantitative manner.

Filling Losses. The filling loss prediction is more reliable than that for breathing loss. The basic cause of loss is the displacement of the air-vapor mixture by the incoming liquid. Once again, the experience varies with the company and the location. The recommended equation is

$$F = \left(\frac{PV}{A}\right)(K_f) \tag{5.22}$$

Where:		Metric	English
	P = true vapor pressure	kPa(g)	psig
	V = volume of liquid in	m^3	bbl
	F = filling loss	m^3	bbl
	A = conversion constant	22 740	3300

The value for K_f is found from the table below.

Tank Turnovers per Year	K_f	
	Refineries	Fields and Terminals
0-10	1.0	1.00
12	0.91	1.00
15	0.75	1.00
20	0.59	1.00
25	0.50	1.00
30	0.47	1.00
40	0.44	1.00
40-60		0.80
60-100		0.50

Conservation type (floating head) tanks are used to reduce losses. The Pan Type floating-head tank is primarily of historical interest. Of the two other common types, pontoon floating roof and double deck, each has its own particular advantages.

Filling losses are usually negligible. To estimate total losses, the following rule-of-thumb may be used.

3.8-4.6 m^3 per meter of diameter per bar of TVP per year

[0.5-0.6 API bbl per foot of diameter per psi of TVP per year]

Additional information on tank storage is available from the American Petroleum Institute.(5.21,5.22)

Automatic Custody Transfer (ACT)

When producing to a pipeline, a continuous monitoring system may be used instead of large tanks that are emptied periodically. Figure 5.10 shows the systems often used.

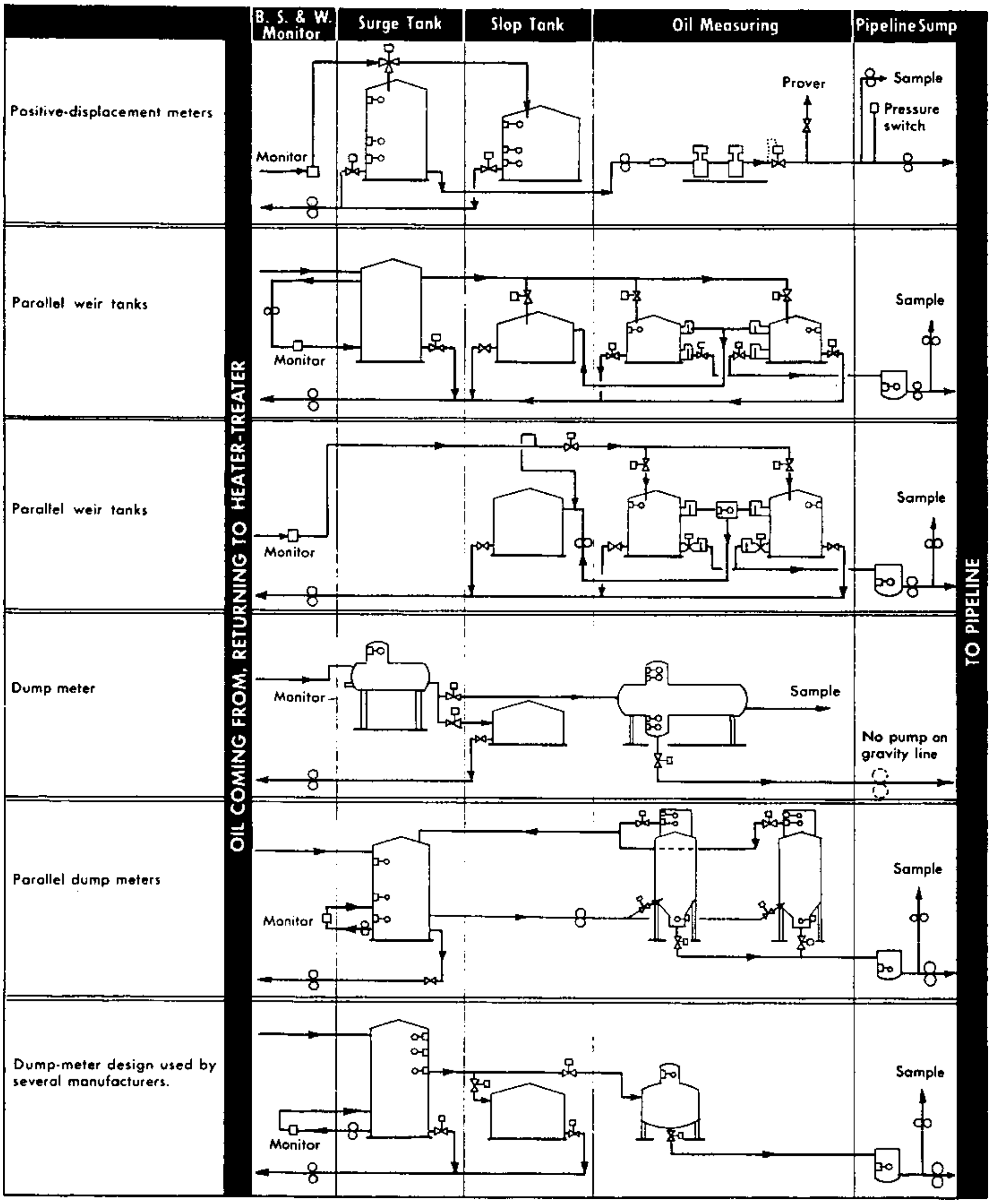


Figure 5.10 Various Types of ACT Systems

The monitor for B.S. & W. (basic sediment and water) rejects any oil not meeting specifications and sends it to the slop tank for retreatment. A surge tank may, or may not, be used depending on circumstances.

The most common method of oil metering is the positive displacement (p.d.) meter. A weir tank is used, particularly if tanks are already on site. Weirs are added, top and bottom, to control the volume passed on each fill and emptying cycle. The dump (metering) tank is like any metering vessel using level controls.

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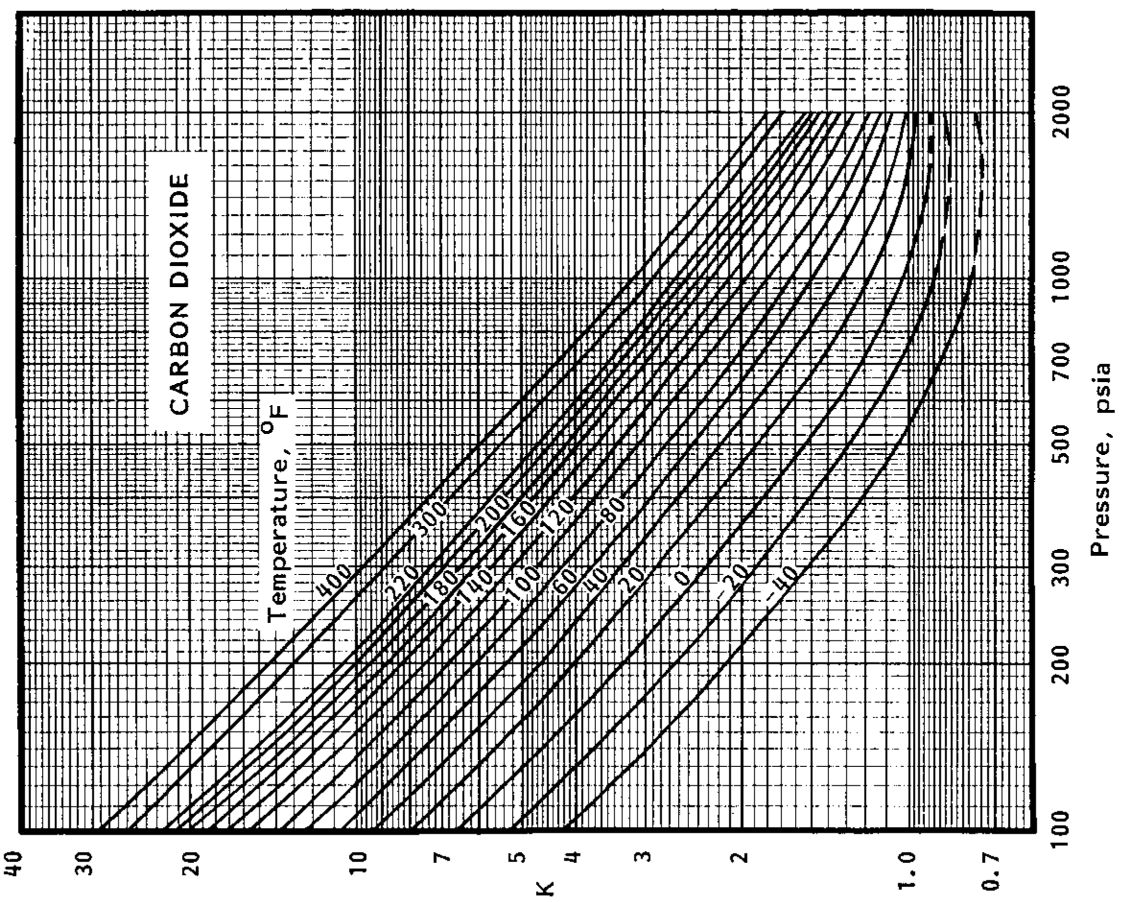
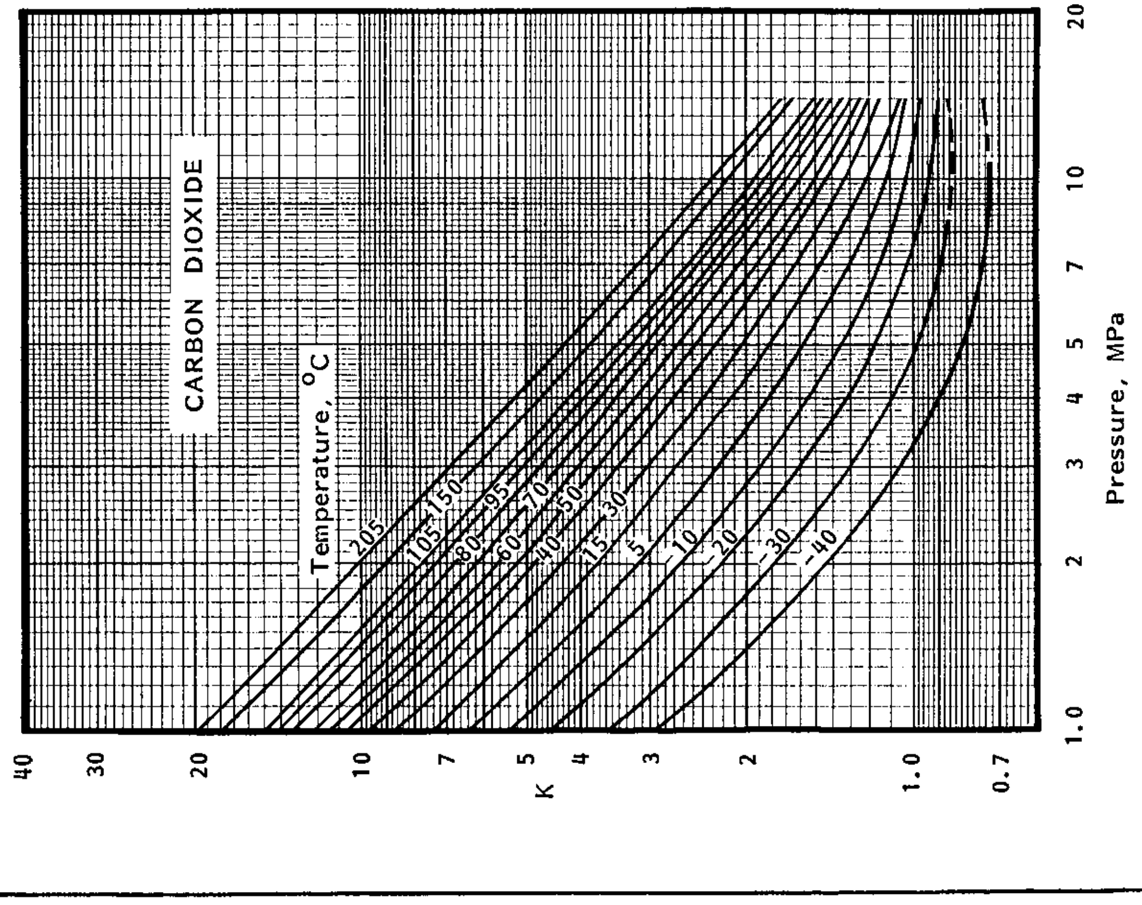
APPENDIX 5A

This appendix contains K value charts for Carbon Dioxide, Hydrogen Sulfide, Nitrogen and the paraffin hydrocarbons methane through decane. Both metric and English unit charts are included.

They are used in all examples involving K values. They also are compatible with the computer programs that go with this book.


CARBON DIOXIDE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
-40
K
40
30
20
10
7
5
4
3
2
1.0
0.7
Pressure, psia
100
200
300
500
700
1000
2000


CARBON DIOXIDE
Temperature, °C
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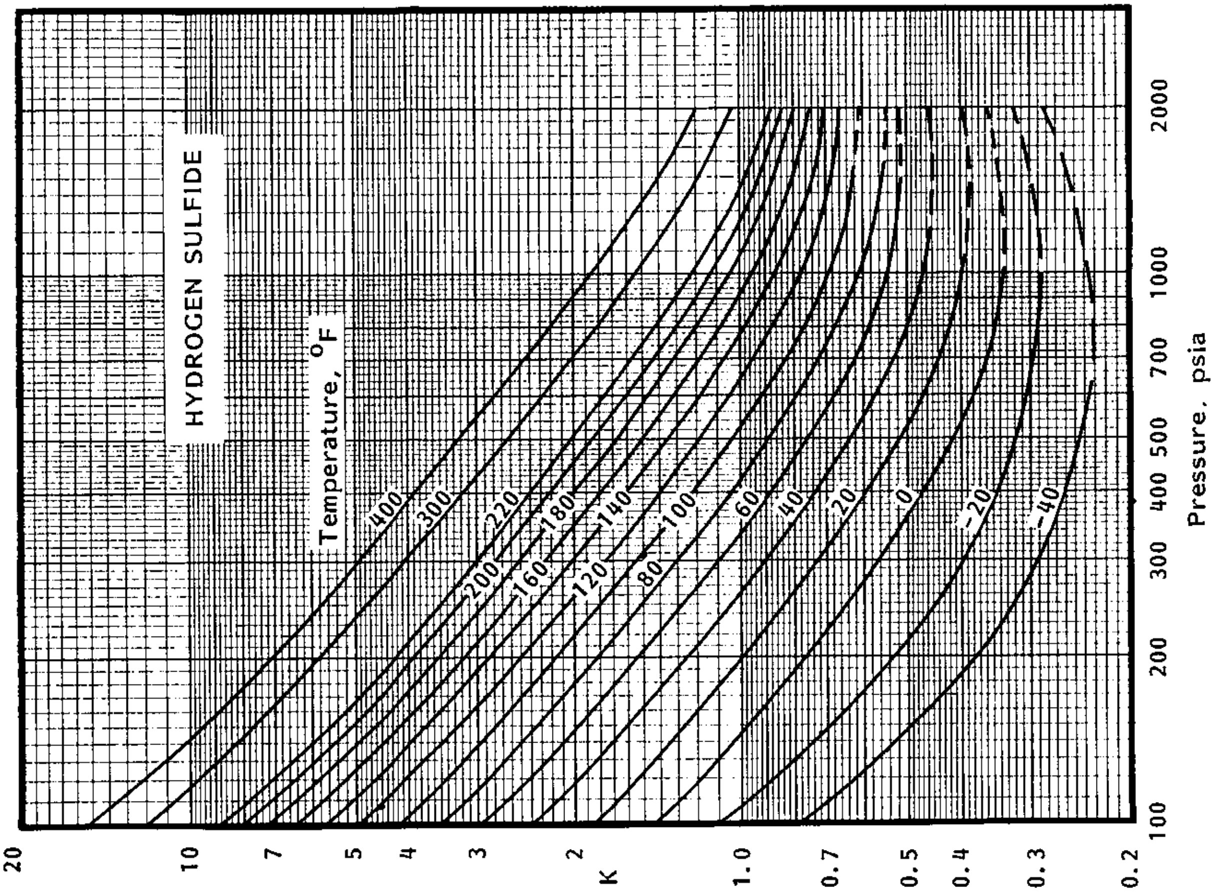
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
-40
K
40
30
20
10
7
5
4
3
2
1.0
0.7
Pressure, MPa
1.0
2
3
4
5
7
10
20


HYDROGEN SULFIDE
Temperature, °F
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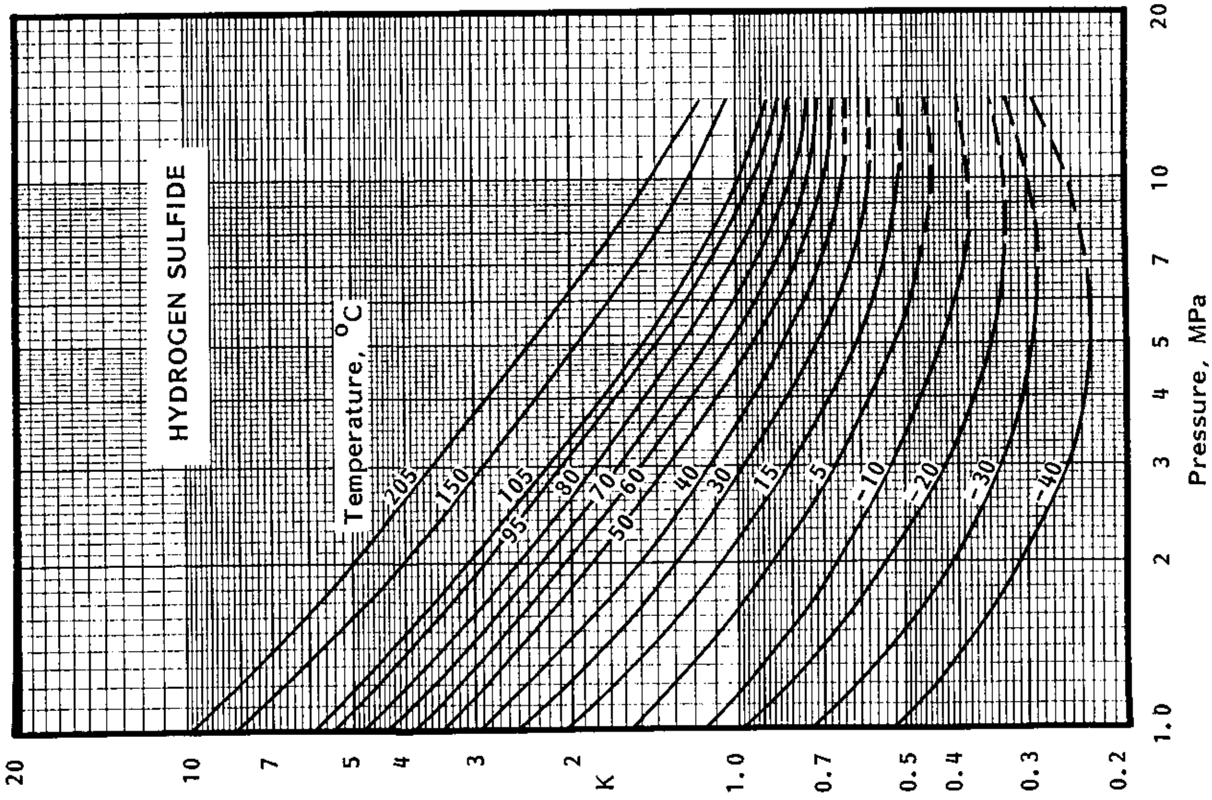
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
-40
K
20
10
7
5
4
3
2
1.0
0.7
0.5
0.4
0.3
0.2
Pressure, psia
100
200
300
400
500
700
1000
2000


HYDROGEN SULFIDE
Temperature, °C
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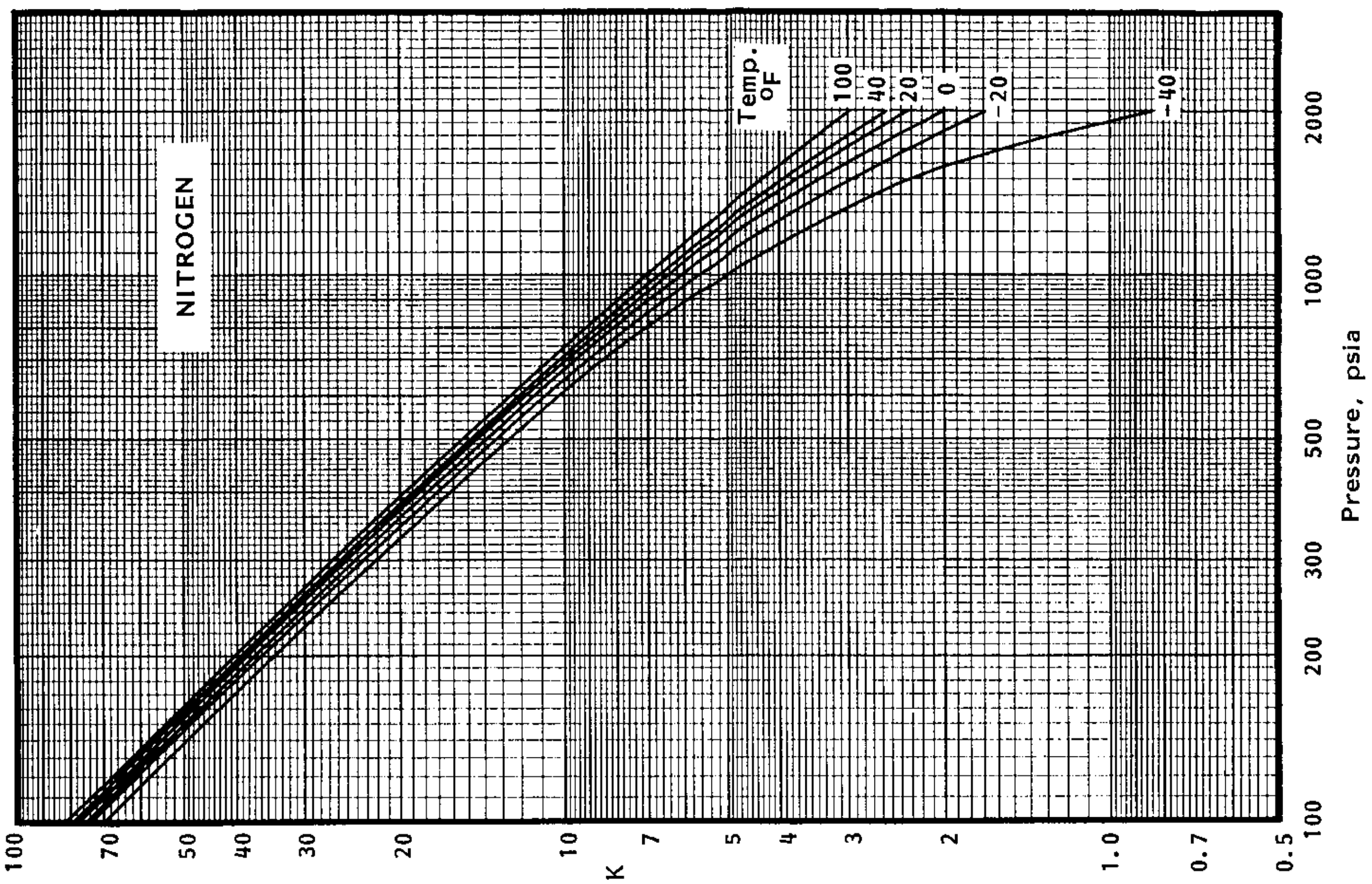
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
-40
K
20
10
7
5
4
3
2
1.0
0.7
0.5
0.4
0.3
0.2
Pressure, MPa
1.0
2
3
4
5
7
10
20


NITROGEN
Temp. °F
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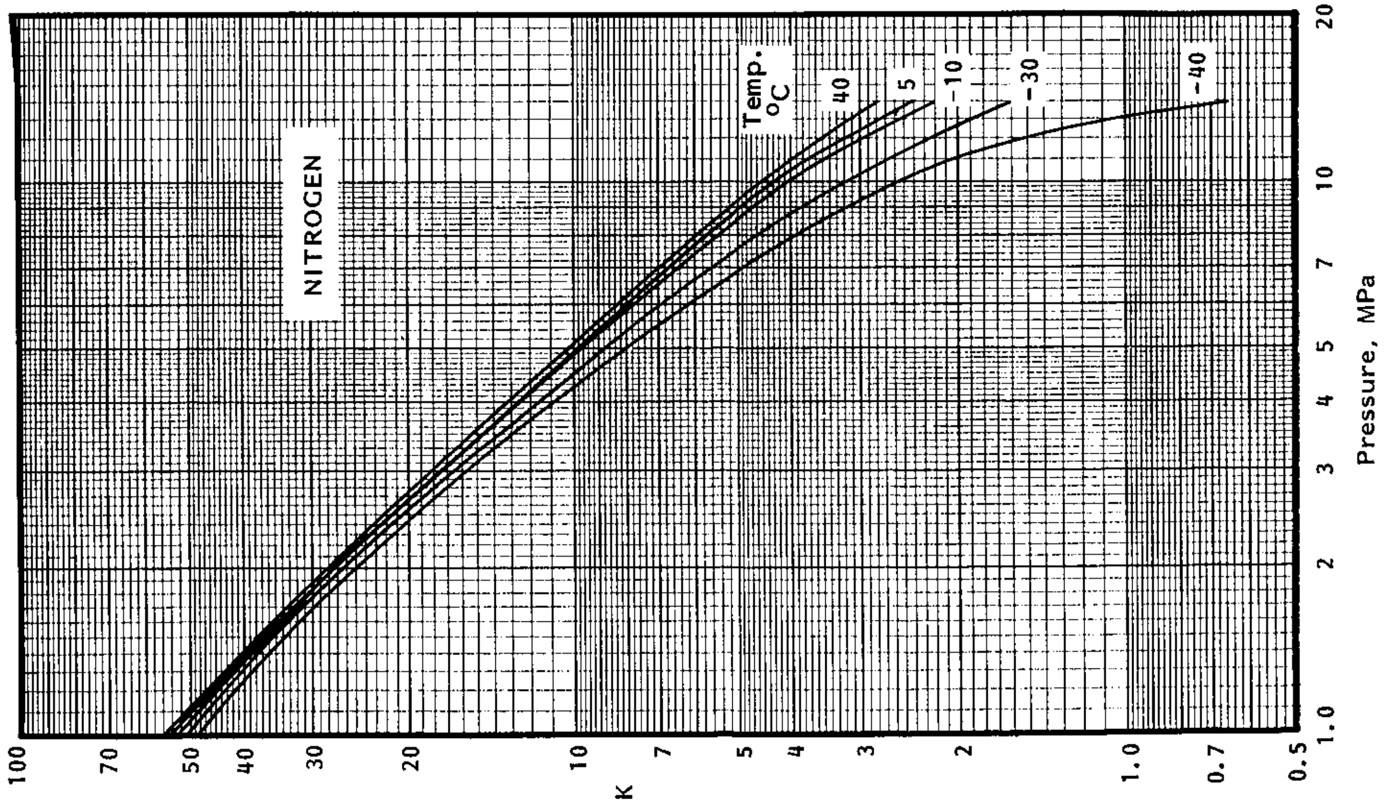
20
0
-20
-40
K
100
70
50
40
30
20
10
7
5
4
3
2
1.0
0.7
0.5
Pressure, psia
100
200
300
500
1000
2000


NITROGEN
Temp. °C
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5
-10
-30
-40
K
100
70
50
40
30
20
10
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5
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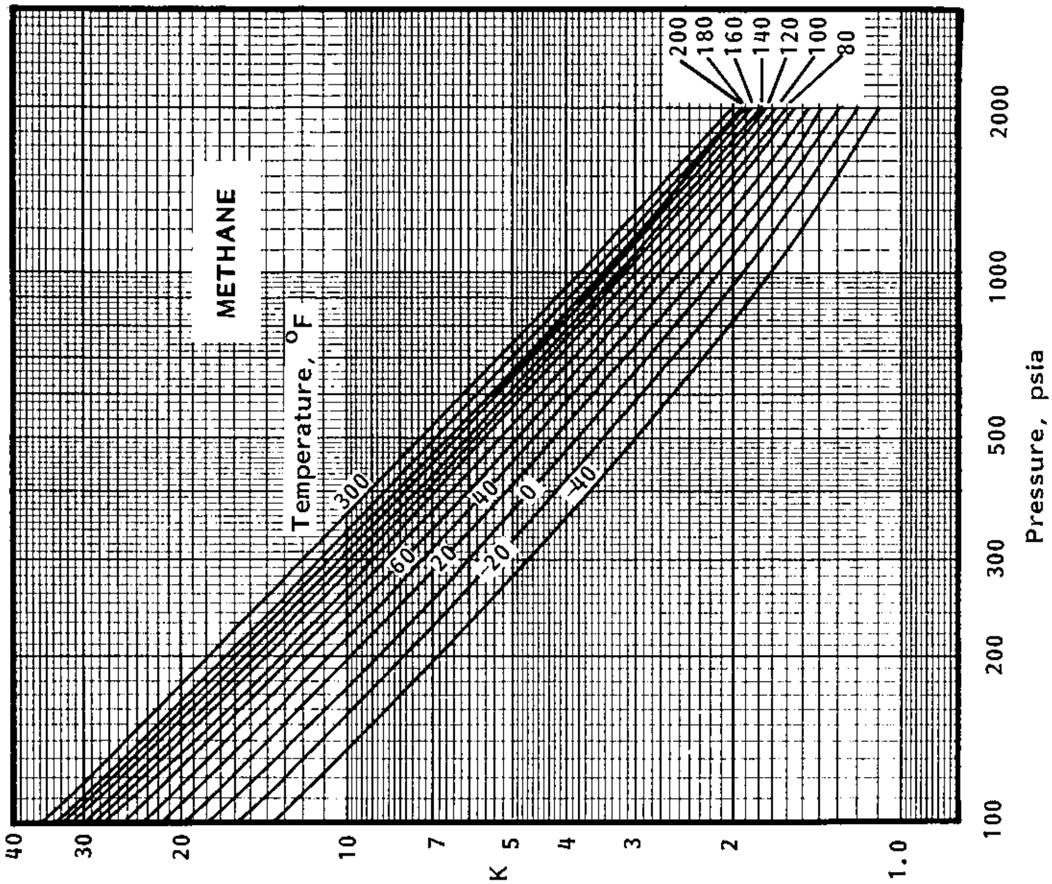
Pressure, MPa
1.0
2
3
4
5
7
10
20


METHANE
Temperature, °F
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-20
-40
200
180
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140
120
100
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K
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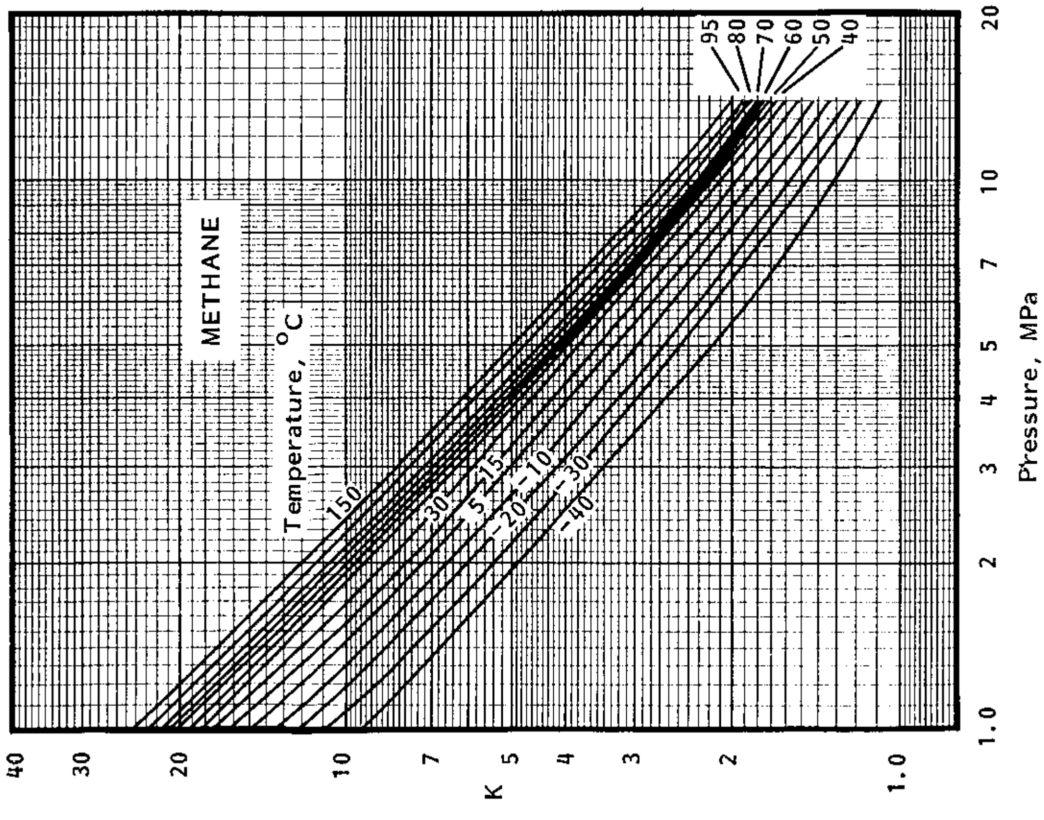
3
2
1.0
Pressure, psia
100
200
300
500
1000
2000


METHANE
Temperature, °C
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-10
-20
-30
-40
95
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70
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50
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K
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1.0
Pressure, MPa
1.0
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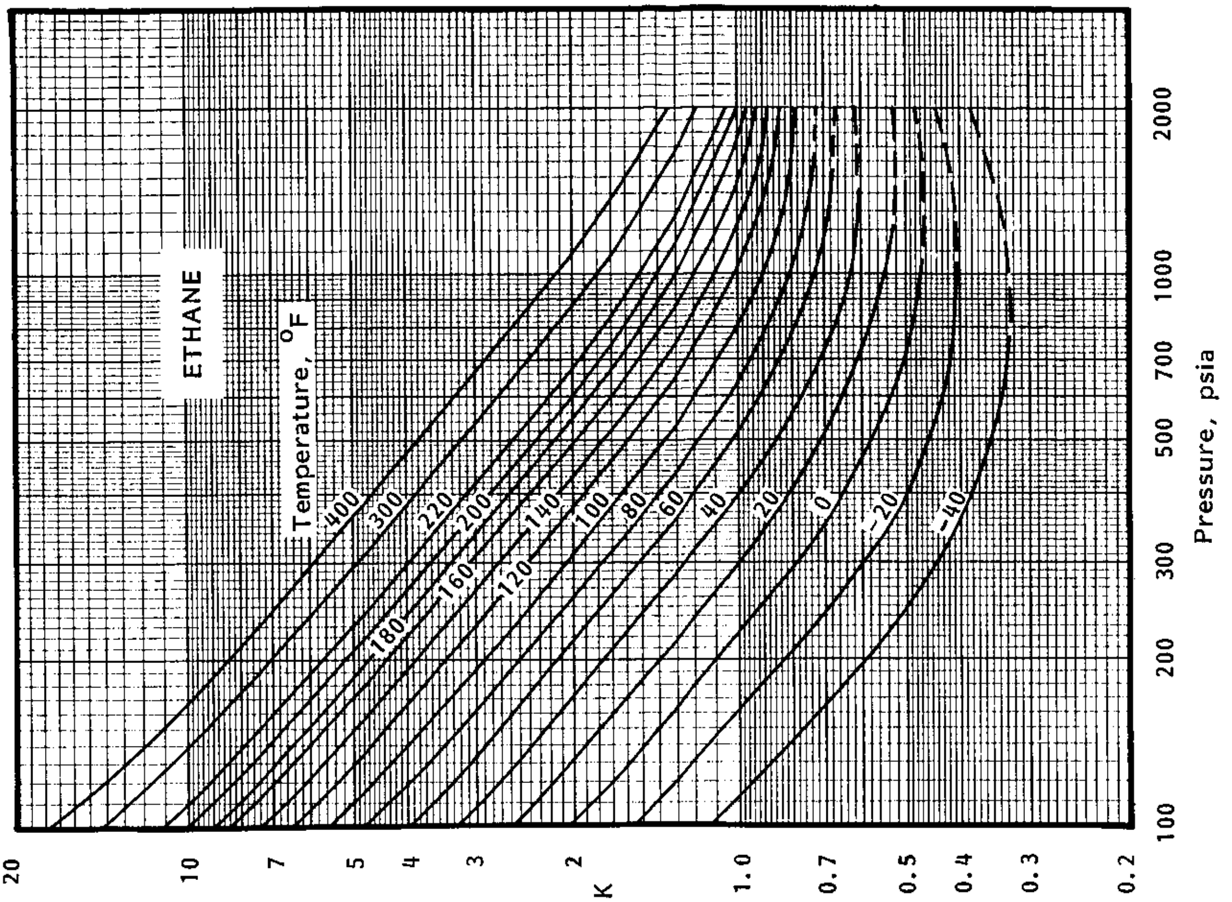
7
10
20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Temperature, °F
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-40
K
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4
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0.4
0.3
0.2
Pressure, psia
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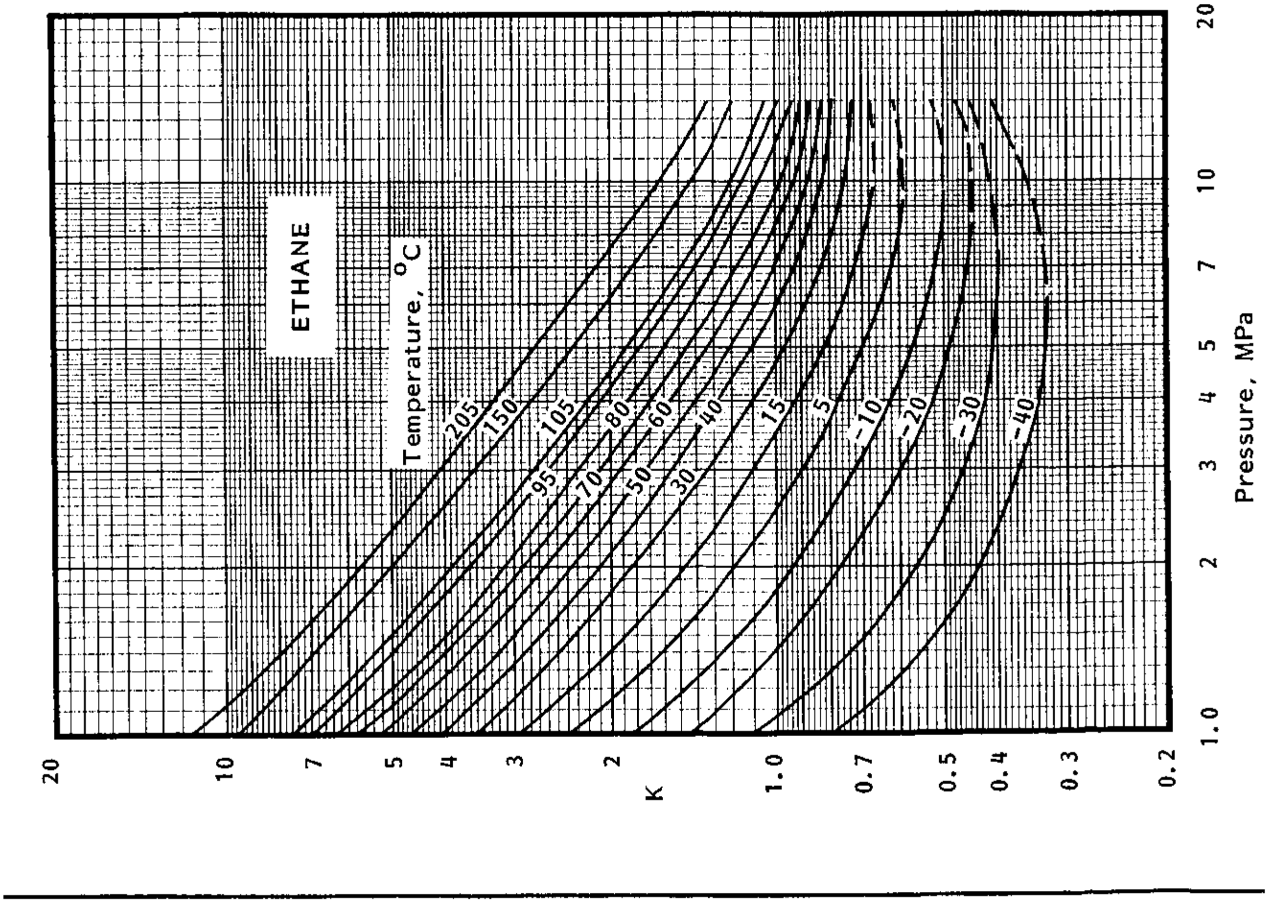
500
700
1000
2000


ETHANE
Temperature, °C
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-20
-30
-40
K
20
10
7
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4
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1.0
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0.3
0.2
Pressure, MPa
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10
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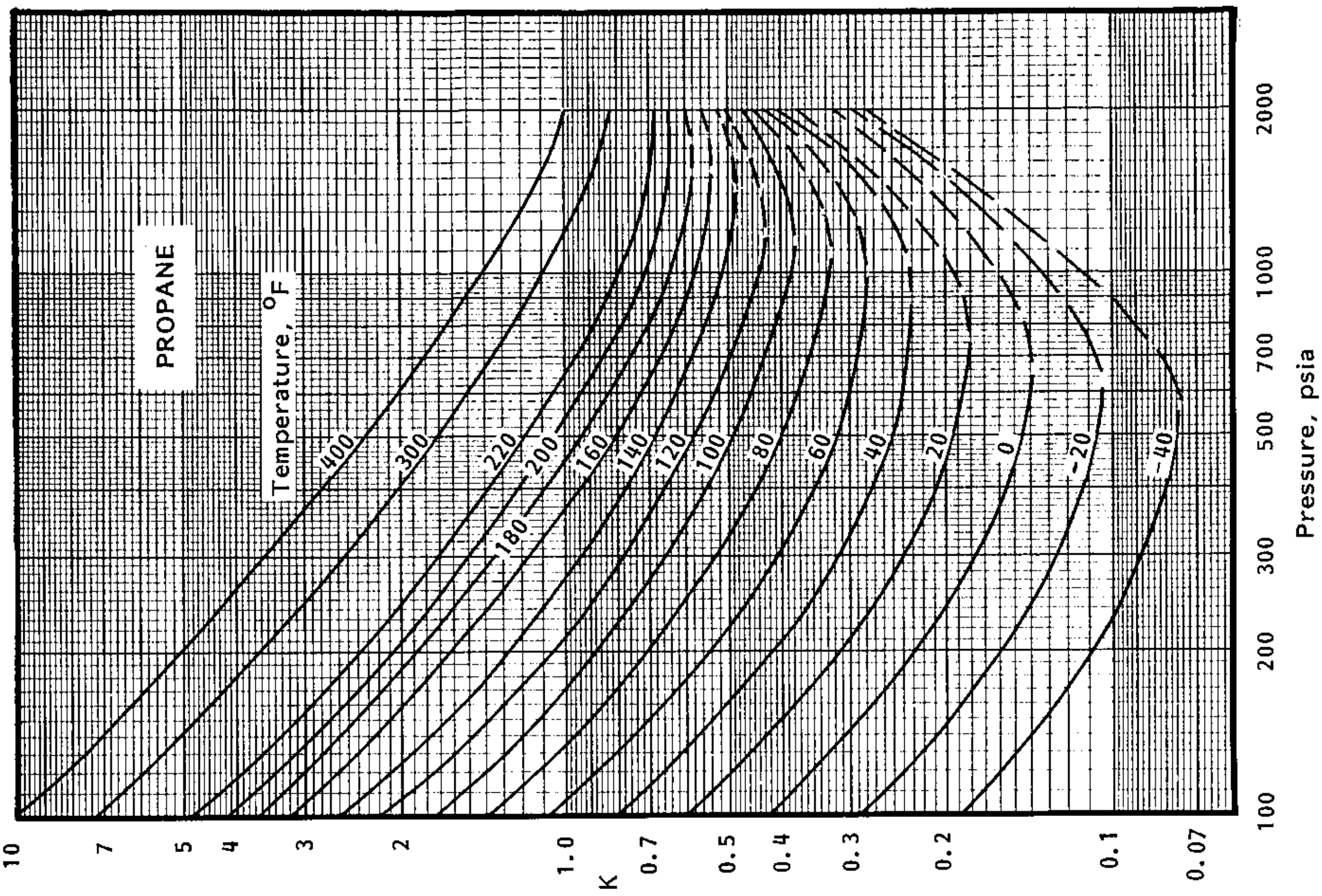
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-20
-40
K
10
7
5
4
3
2
1.0
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0.3
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0.07
Pressure, psia
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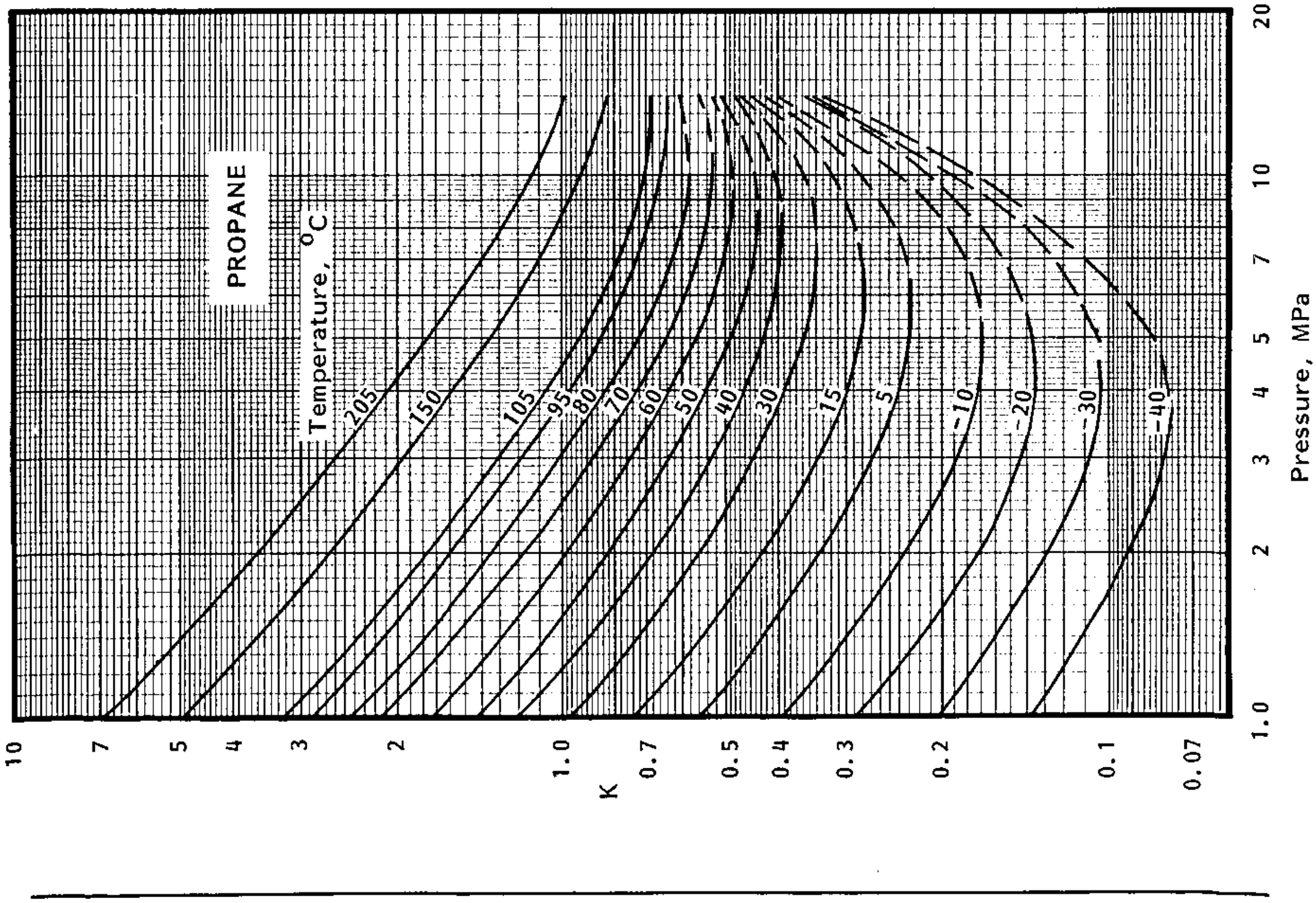
300
500
700
1000
2000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-20
-30
-40
K
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7
5
4
3
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1.0
0.7
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0.1
0.07
Pressure, MPa
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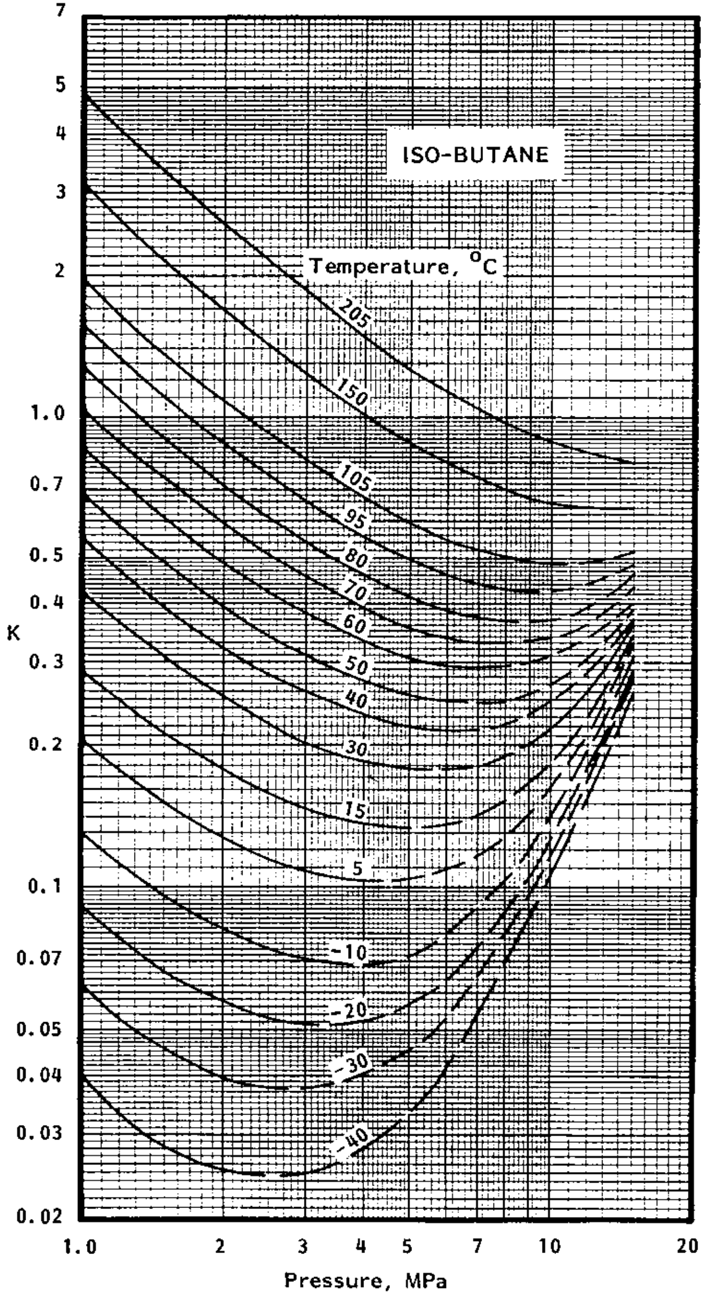
2
3
4
5
7
10
20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15
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-10
-20
-30
-40
K
7
5
4
3
2
1.0
0.7
0.5
0.4
0.3
0.2
0.1
0.07
0.05
0.04
0.03
0.02
Pressure, MPa
1.0
2
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4
5
7
10
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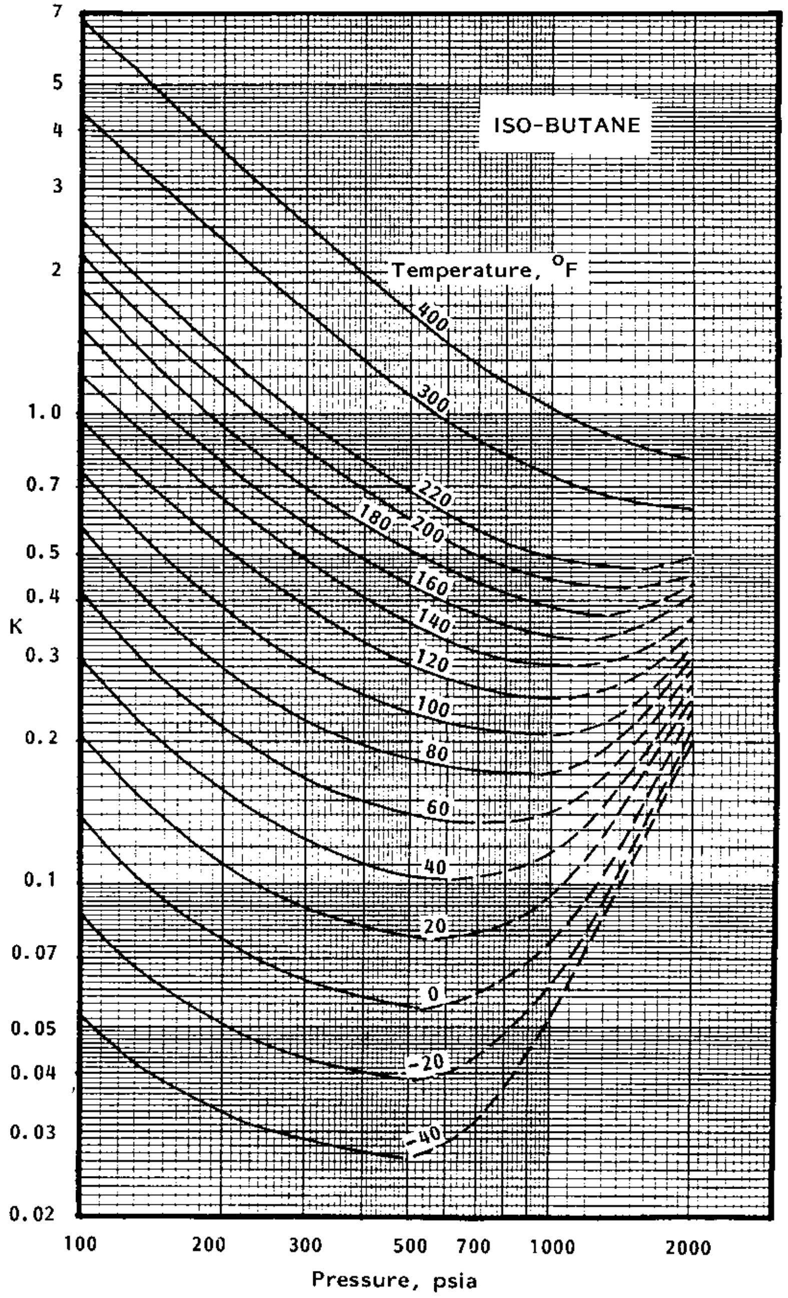
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K
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0.5
0.4
0.3
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0.04
0.03
0.02
Pressure, psia
100
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1000
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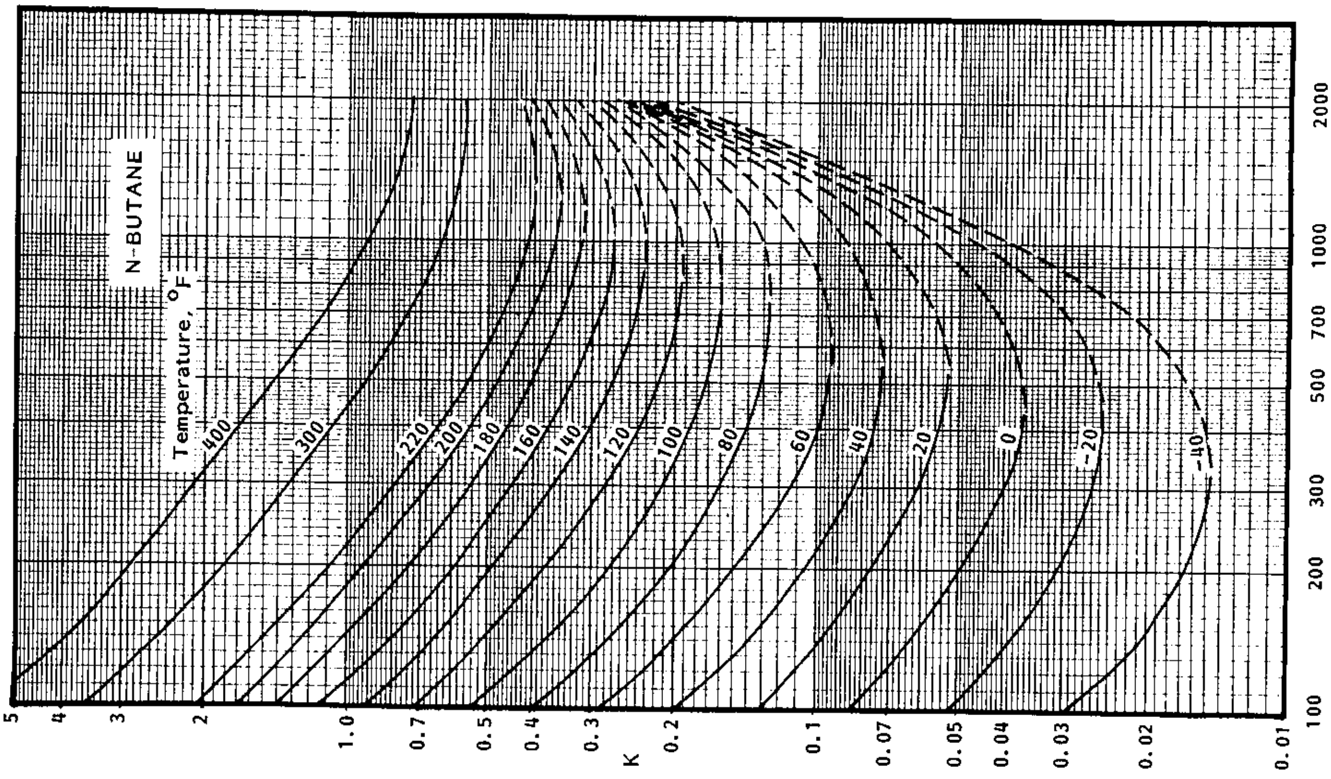
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-40
K
5
4
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1.0
0.7
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0.4
0.3
0.2
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Pressure, psia
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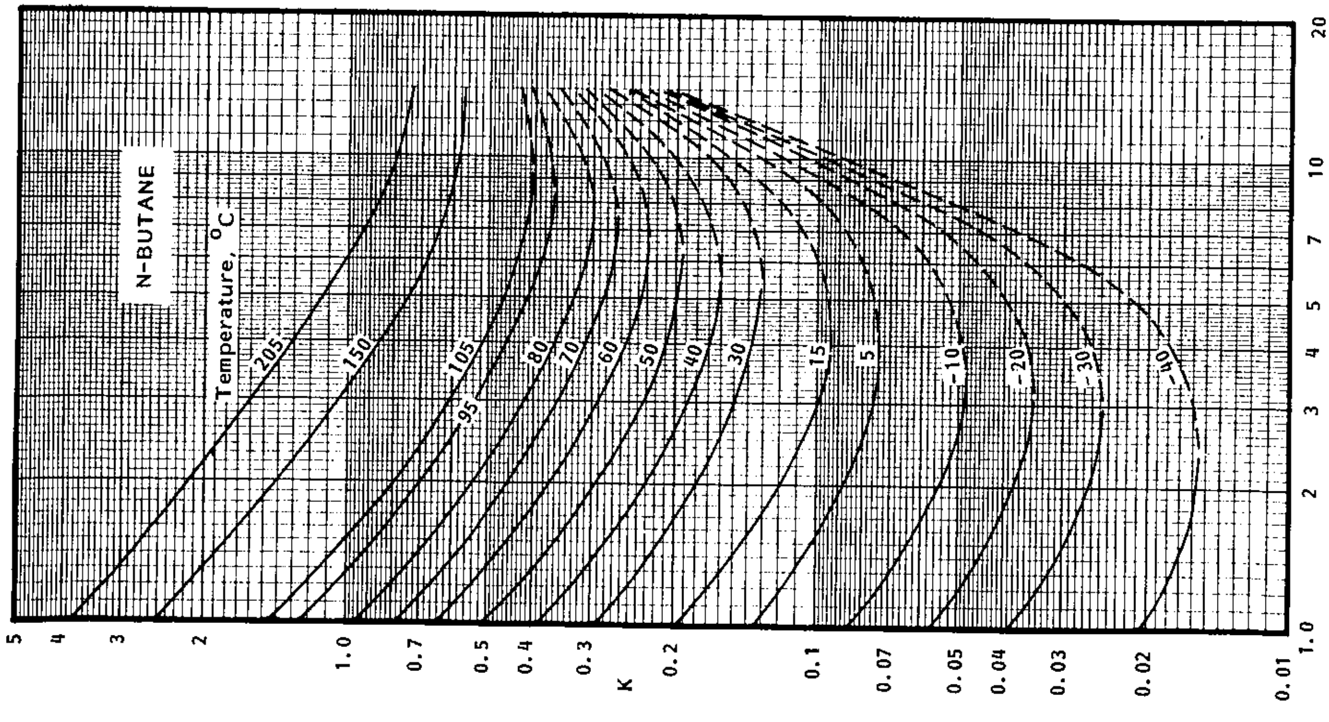
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-40
K
5
4
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1.0
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0.5
0.4
0.3
0.2
0.1
0.07
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0.01
Pressure, MPa
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4
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7
10
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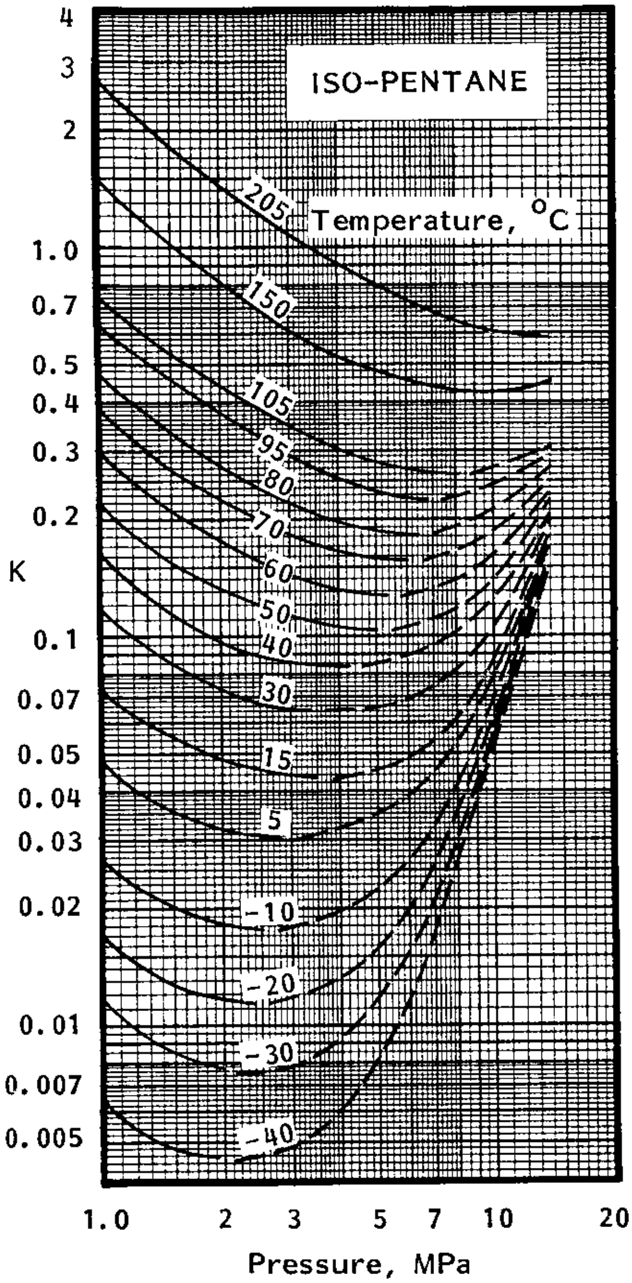
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Temperature, °C
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-30
-40
K
4
3
2
1.0
0.7
0.5
0.4
0.3
0.2
0.1
0.07
0.05
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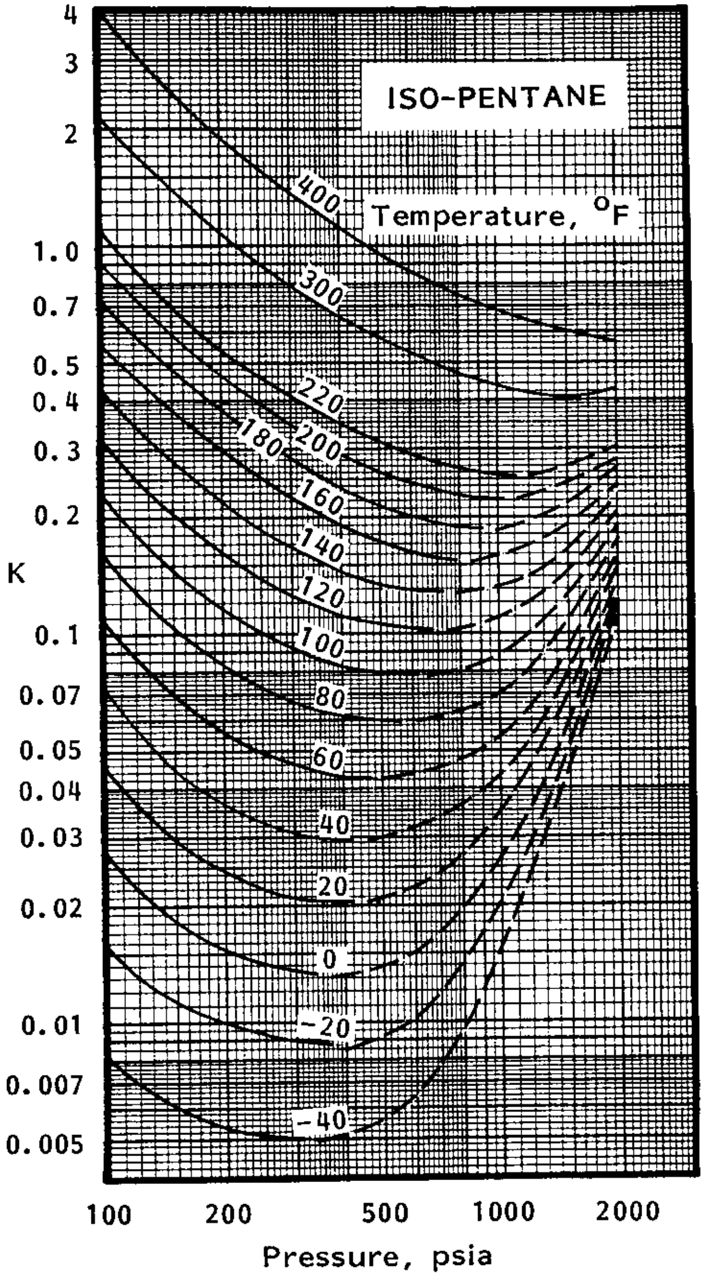
Pressure, MPa


ISO-PENTANE
Temperature, °F
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300
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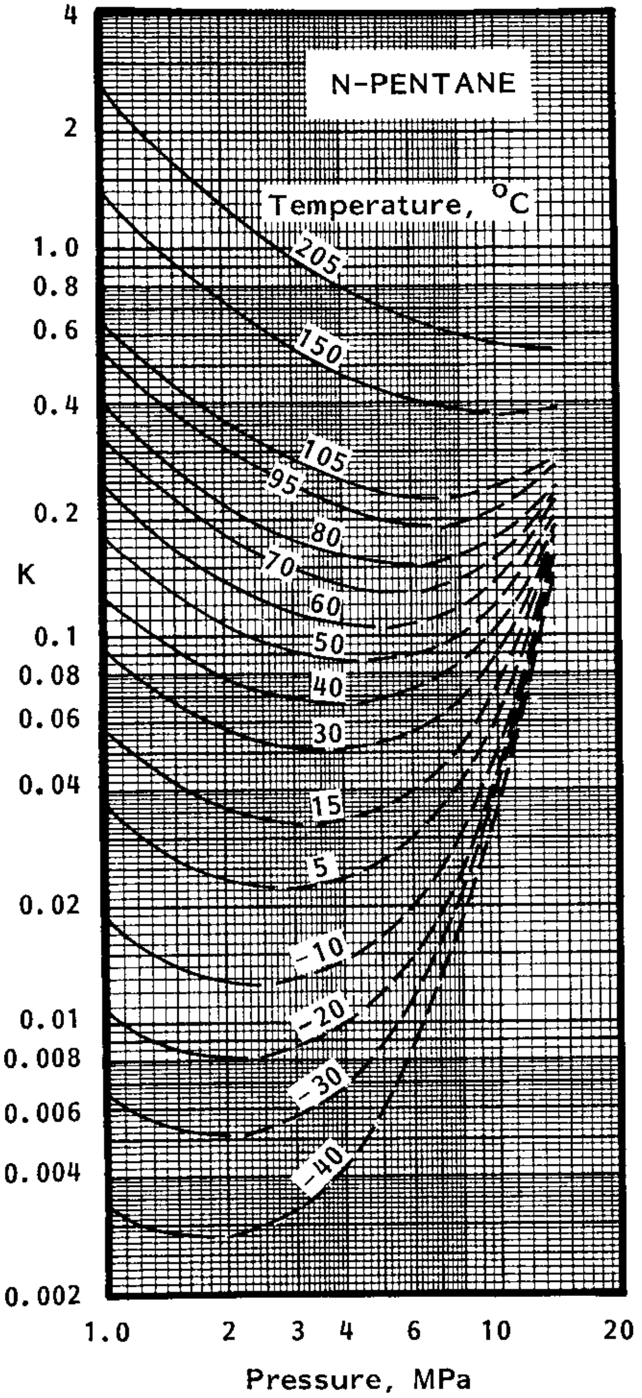
140
120
100
80
60
40
20
0
-20
-40
K
4
3
2
1.0
0.7
0.5
0.4
0.3
0.2
0.1
0.07
0.05
0.04
0.03
0.02
0.01
0.007
0.005
100
200
500
1000
2000
Pressure, psia


N-PENTANE
Temperature, °C
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15
5
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-20
-30
-40
K
4
2
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
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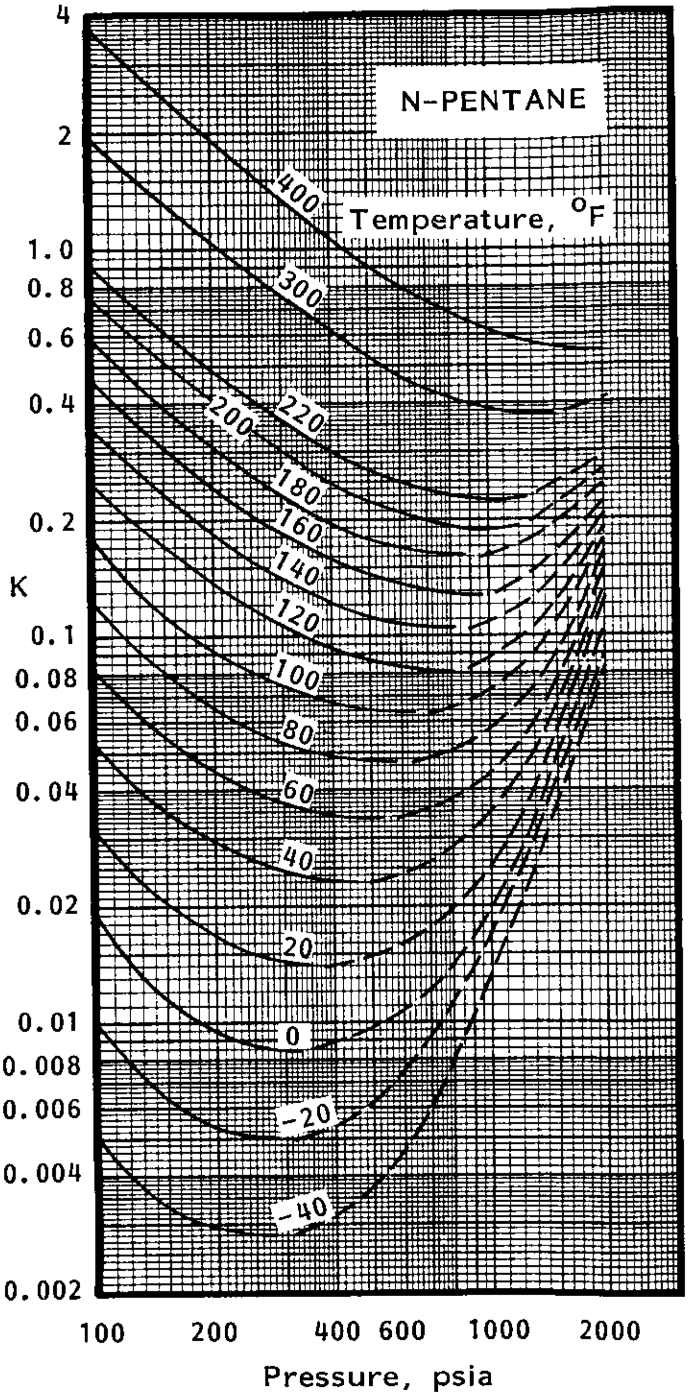
4
6
10
20
Pressure, MPa


N-PENTANE
Temperature, °F
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140
120
100
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60
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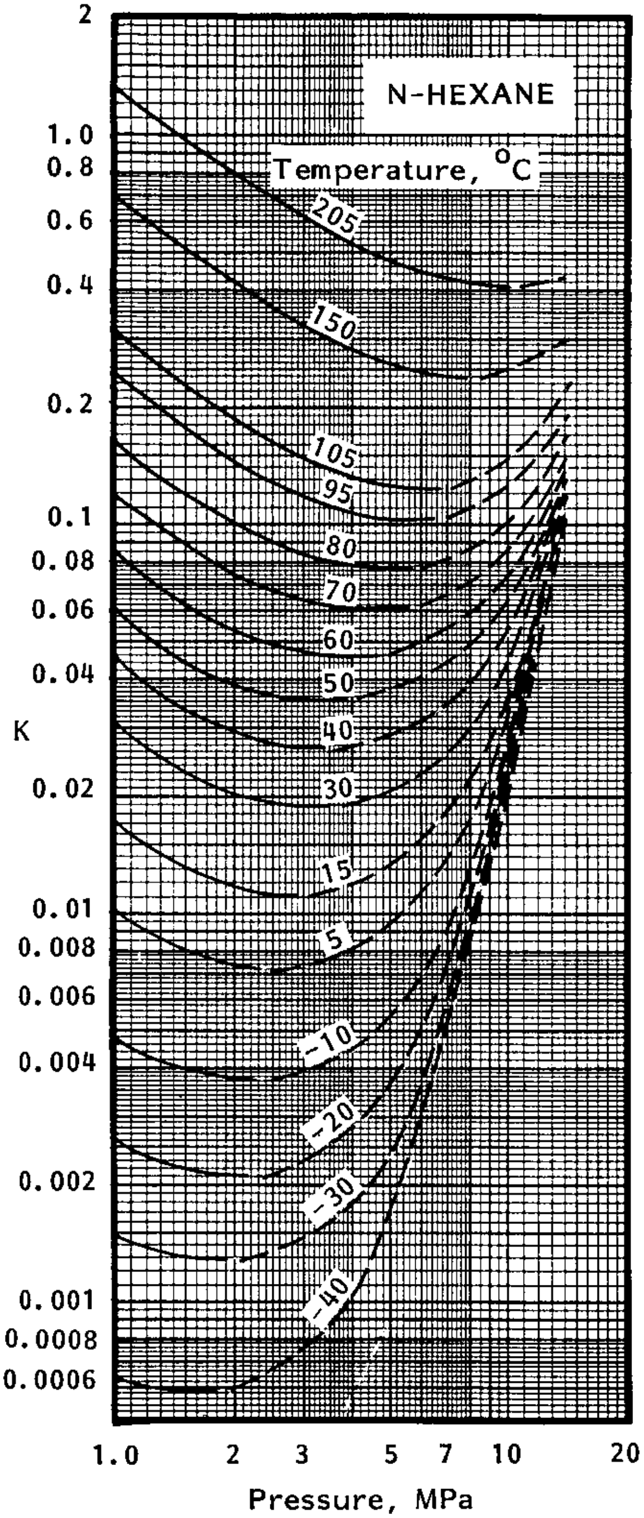
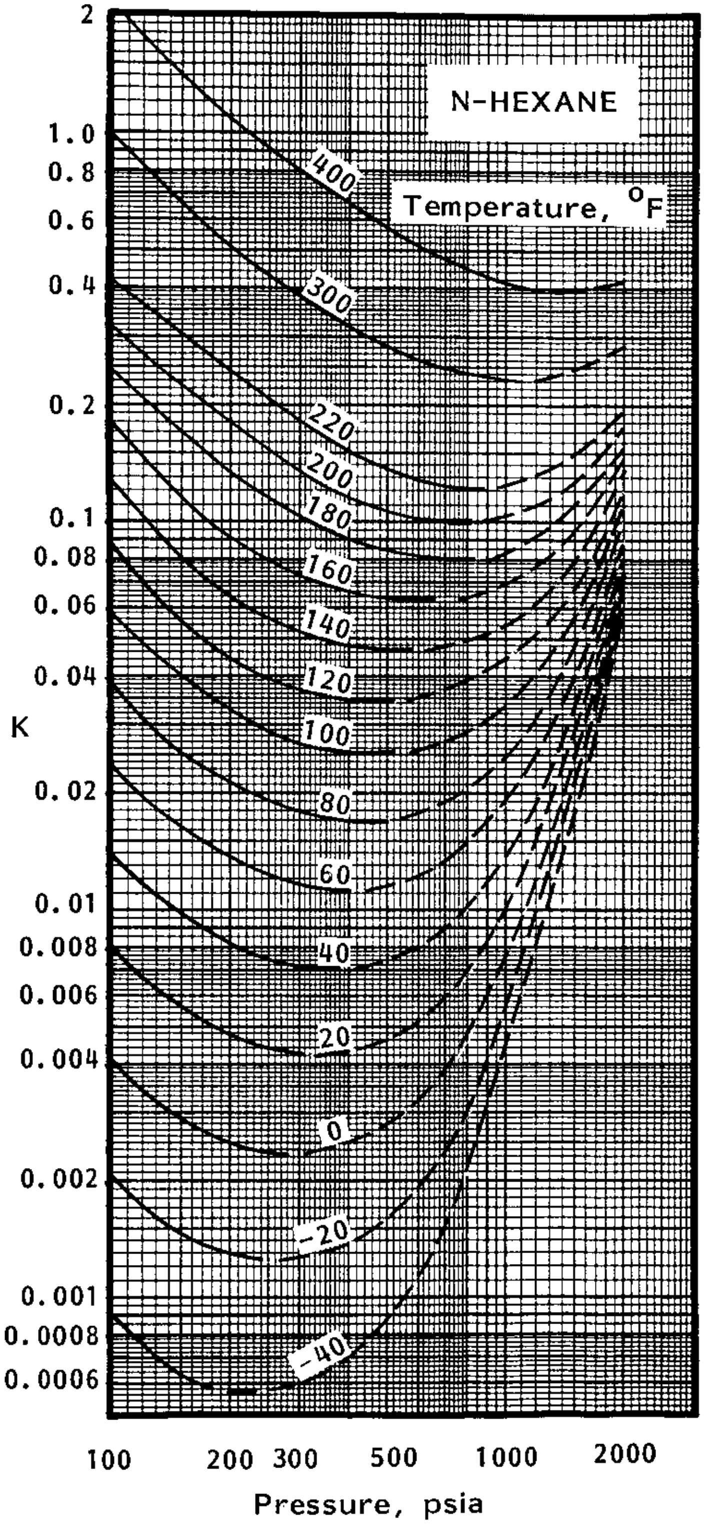
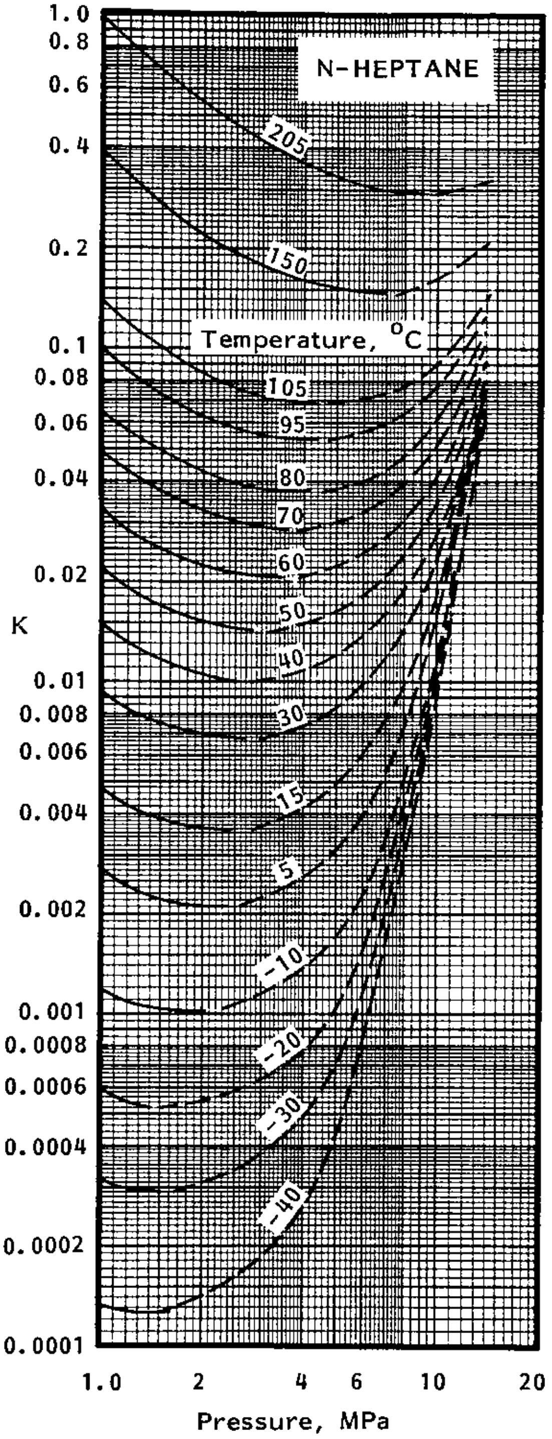
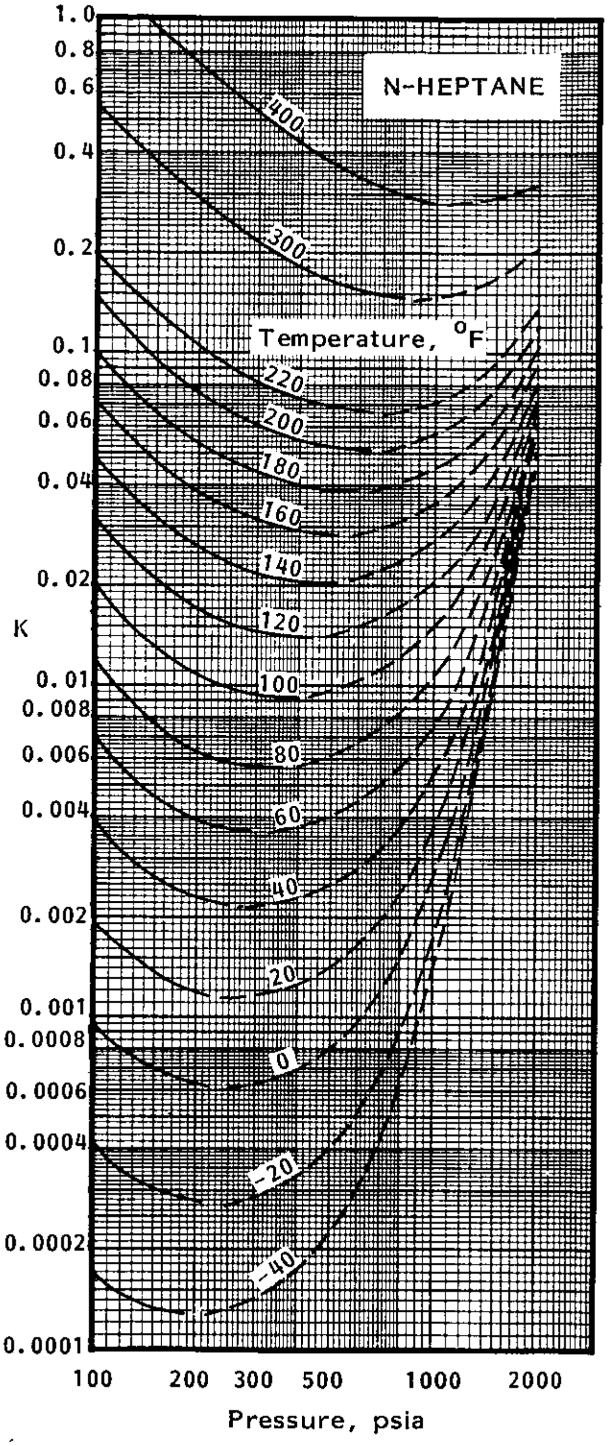
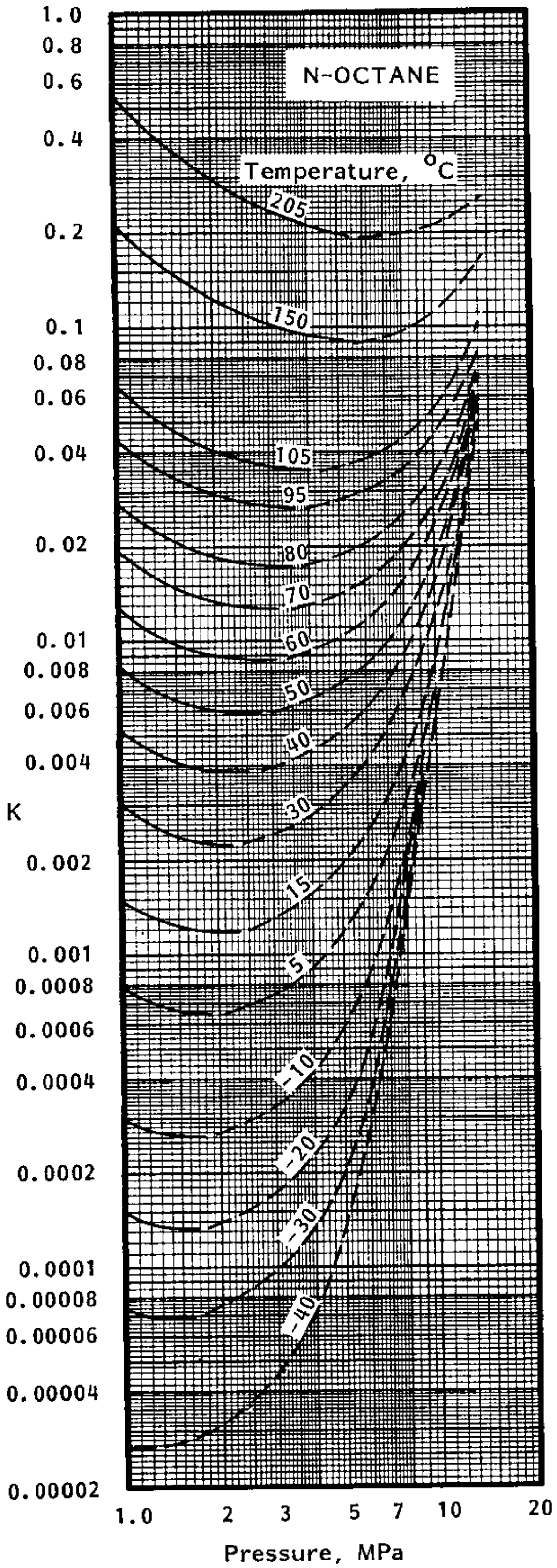
20
0
-20
-40
K
4
2
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
100
200
400
600
1000
2000
Pressure, psia

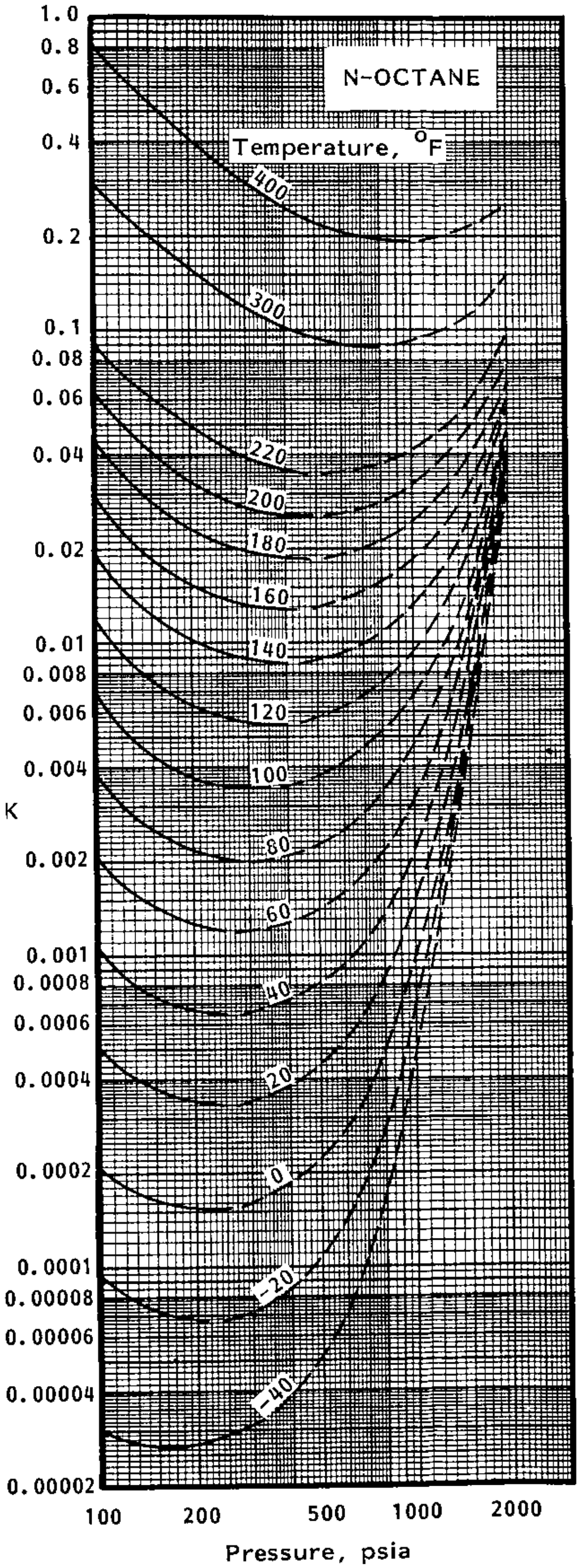

N-HEXANE
Temperature, °C
205
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
-40
2
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
K
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
1.0
2
3
5
7
10
20
Pressure, MPa

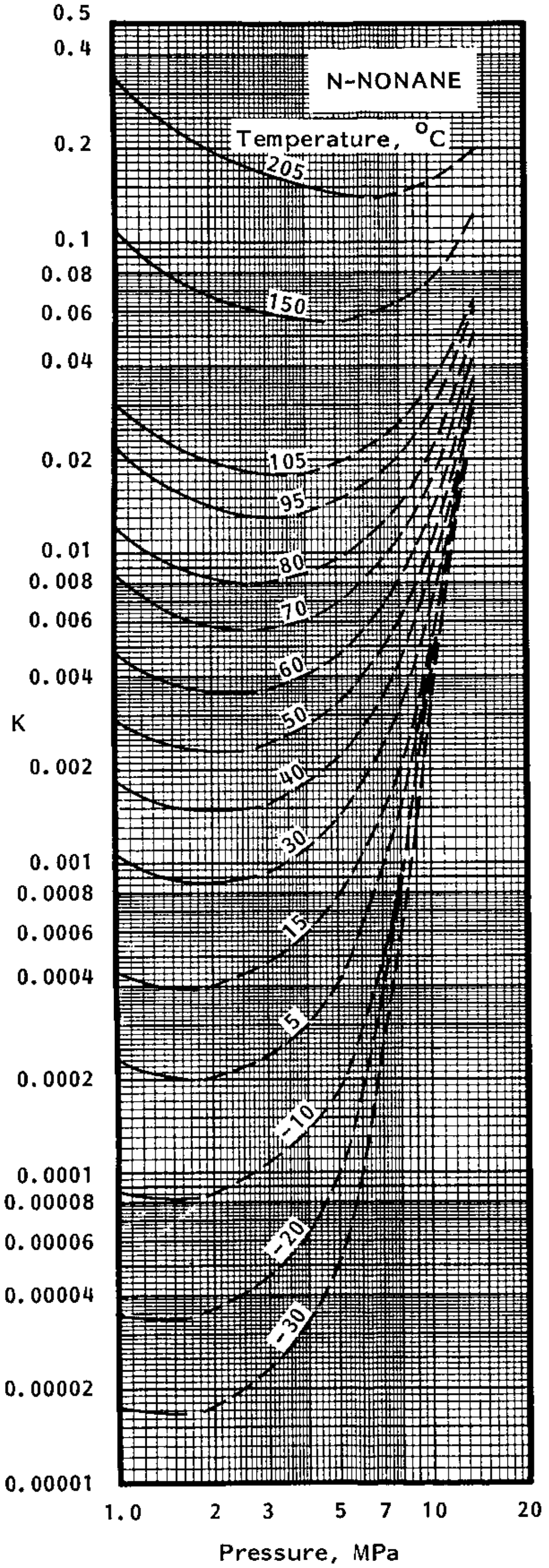

N-HEXANE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
-40
2
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
K
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
100
200
300
500
1000
2000
Pressure, psia

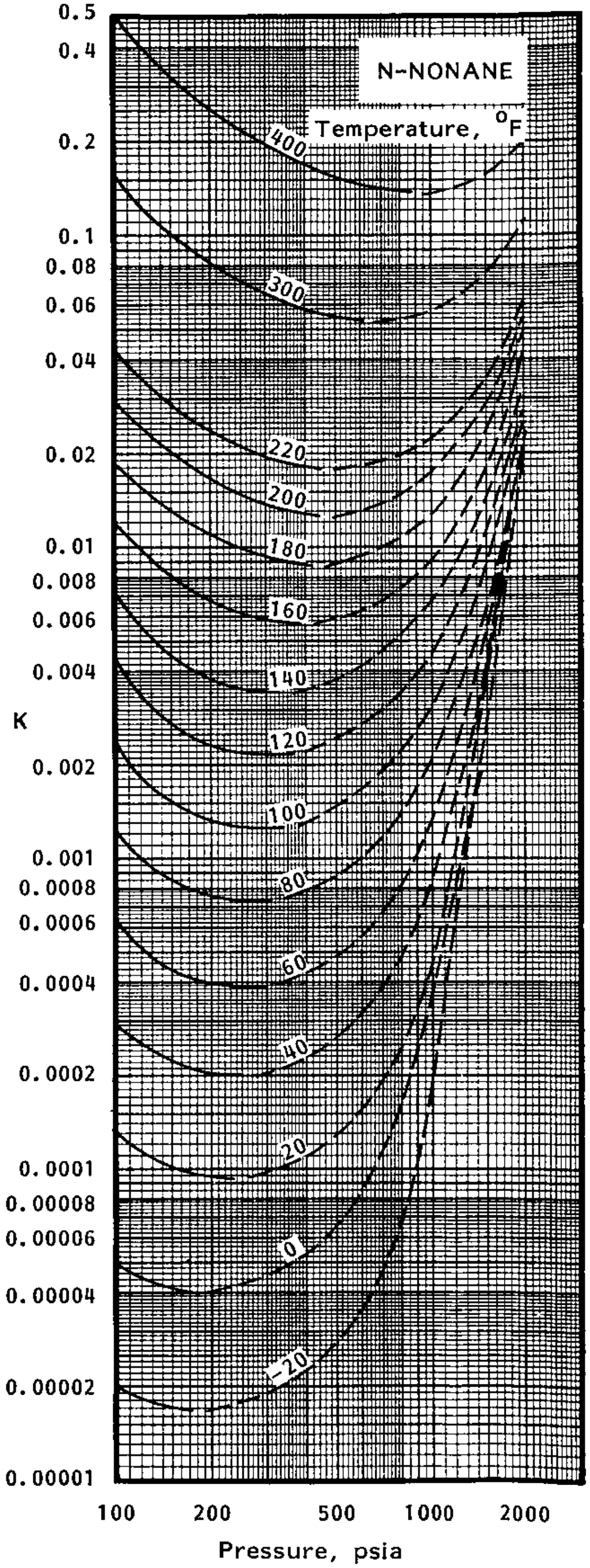

N-HEPTANE
Temperature, °C
205
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
-40
K
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
1.0
2
4
6
10
20
Pressure, MPa

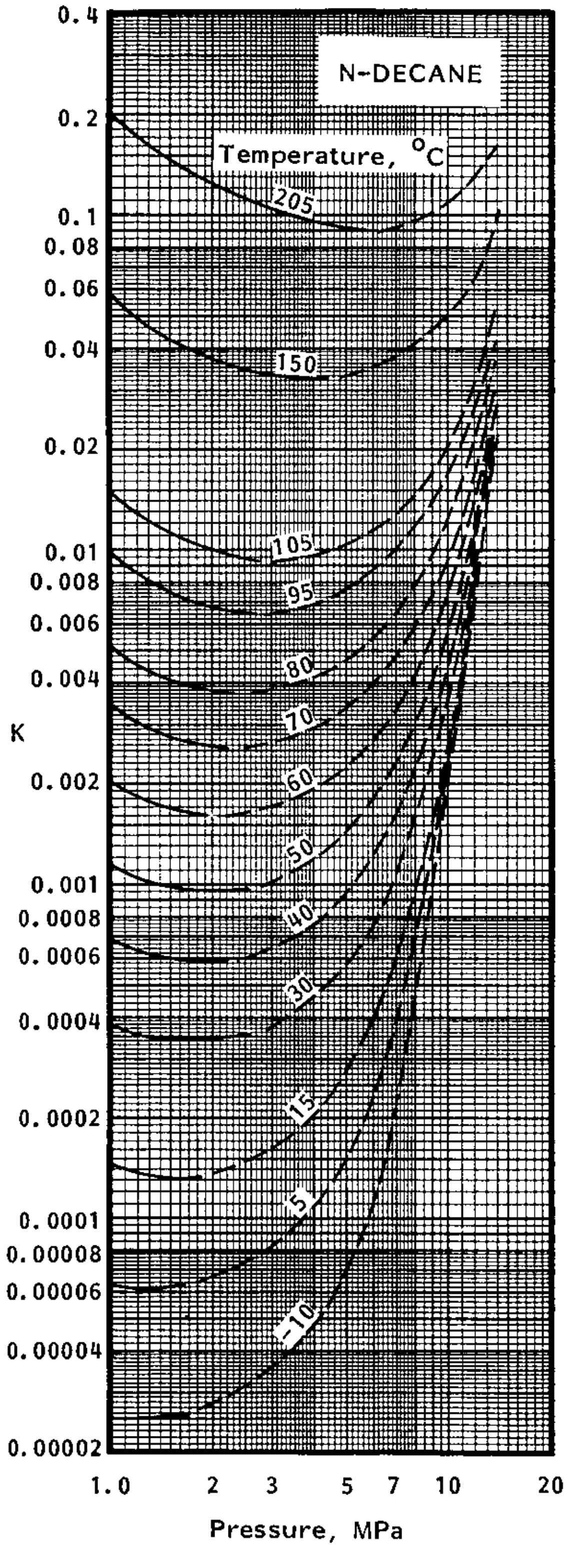

N-HEPTANE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
-40
K
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
100
200
300
500
1000
2000
Pressure, psia

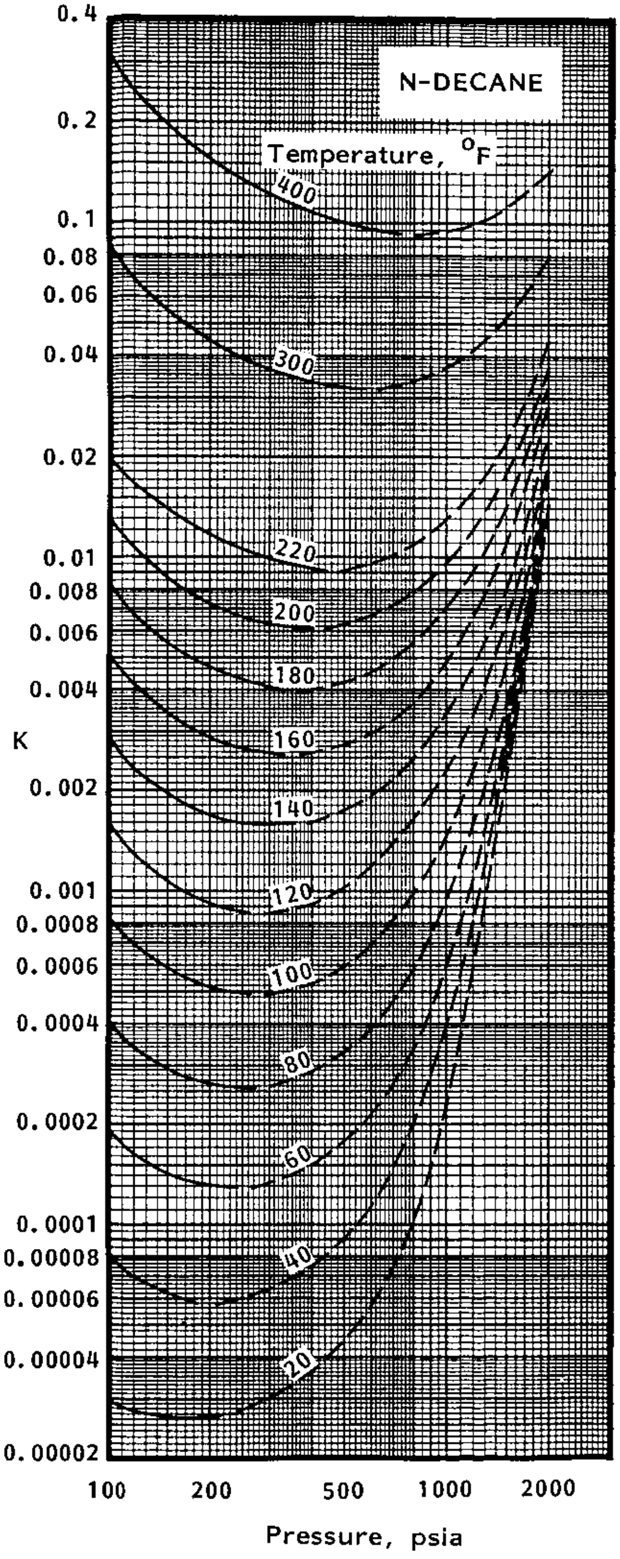

N-OCTANE
Temperature, °C
205
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
-40
K
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
1.0
2
3
5
7
10
20
Pressure, MPa


N-OCTANE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
-40
K
1.0
0.8
0.6
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
100
200
500
1000
2000
Pressure, psia


N-NONANE
Temperature, °C
205
150
105
95
80
70
60
50
40
30
15
5
-10
-20
-30
K
0.5
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
0.00001
1.0
2
3
5
7
10
20
Pressure, MPa


N-NONANE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
0
-20
K
0.5
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
0.00001
100
200
500
1000
2000
Pressure, psia


N-DECANE
Temperature, °C
205
150
105
95
80
70
60
50
40
30
15
5
-10
K
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
1.0
2
3
5
7
10
20
Pressure, MPa


N-DECANE
Temperature, °F
400
300
220
200
180
160
140
120
100
80
60
40
20
K
0.4
0.2
0.1
0.08
0.06
0.04
0.02
0.01
0.008
0.006
0.004
0.002
0.001
0.0008
0.0006
0.0004
0.0002
0.0001
0.00008
0.00006
0.00004
0.00002
100
200
500
1000
2000
Pressure, psia

APPENDIX 5B

The composite charts are based on the work of Latham and Campbell. (Reference 5.2) They are designed primarily for use with crude oils; whereas, Appendix 5A is based on natural gas liquids.

COMPOSITE K VALUES FOR RESERVOIR FLUIDS

COMPONENT	TEMP., °F	PRESSURE, psia									
		100	200	300	400	500	600	700	800	900	1000
METHANE	−20	16.0	8.08	5.50	4.23	3.55	3.09	2.75	2.50	2.32	2.19
	0	17.8	9.06	6.18	4.72	3.89	3.34	2.95	2.68	2.44	2.28
	20	20.3	10.6	7.28	5.59	4.59	3.90	3.43	3.19	2.81	2.58
	40	23.1	11.8	7.95	6.16	5.04	4.30	3.70	3.30	3.00	2.72
	60	25.0	12.8	8.7	6.62	5.42	4.60	4.03	3.60	3.28	2.99
	80	26.9	14.3	9.9	7.78	6.44	5.50	4.81	4.38	3.90	3.65
	100	28.7	14.5	10.0	7.90	6.60	5.60	4.90	4.40	4.00	3.70
	120	33.0	16.5	11.5	8.95	7.30	6.15	5.25	4.90	4.50	4.10
ETHANE	−20	1.31	.708	.519	.438	.391	.366	.354	.345	.343	.347
	0	1.95	1.05	.755	.610	.538	.492	.466	.448	.439	.439
	20	2.50	1.32	.939	.768	.657	.590	.560	.539	.522	.518
	40	3.25	1.75	1.28	1.01	.890	.800	.740	.690	.668	.665
	60	3.95	2.12	1.52	1.21	1.05	.940	.862	.808	.770	.750
	80	4.60	2.48	1.78	1.41	1.20	1.08	.985	.918	.863	.821
	100	5.82	3.04	2.19	1.68	1.41	1.24	1.12	1.05	.980	.930
	120	6.40	3.32	2.31	1.81	1.52	1.36	1.22	1.14	1.08	1.01
PROPANE	−20	.269	.149	.111	.093	.083	.077	.076	.076	.079	.082
	0	.443	.244	.178	.144	.128	.116	.112	.112	.114	.119
	20	.600	.328	.245	.204	.179	.166	.159	.157	.157	.159
	40	.840	.458	.342	.289	.258	.241	.230	.223	.220	.220
	60	1.09	.596	.440	.370	.328	.305	.293	.285	.278	.277
	80	1.33	.730	.530	.430	.375	.345	.326	.312	.305	.300
	100	1.75	.930	.670	.550	.480	.430	.408	.385	.373	.365
	120	2.18	1.15	.810	.660	.575	.520	.489	.455	.438	.410
i-BUTANE	−20	.097	.058	.046	.039	.036	.034	.034	.033	.034	.035
	0	.145	.085	.067	.058	.054	.052	.051	.051	.052	.054
	20	.238	.124	.094	.081	.074	.071	.070	.070	.071	.074
	40	.298	.170	.133	.117	.108	.102	.100	.100	.100	.104
	60	.419	.242	.186	.159	.144	.136	.132	.130	.130	.135
	80	.561	.338	.248	.200	.180	.169	.161	.158	.155	.156
	100	.79	.441	.325	.270	.239	.218	.206	.195	.190	.189
	120	.990	.550	.386	.319	.282	.256	.242	.231	.227	.222
n-BUTANE	−20	.058	.035	.028	.024	.022	.021	.021	.022	.024	.026
	0	.090	.053	.041	.036	.033	.032	.032	.033	.034	.037
	20	.145	.082	.064	.054	.049	.046	.045	.046	.047	.050
	40	.203	.125	.097	.086	.079	.076	.074	.074	.074	.076
	60	.295	.171	.132	.116	.107	.101	.097	.096	.096	.098
	80	.418	.233	.178	.149	.136	.129	.125	.121	.120	.120
	100	.55	.311	.261	.194	.176	.162	.155	.150	.148	.148
	120	.750	.415	.305	.249	.221	.203	.189	.182	.180	.178
i-PENTANE	−20	.017	.010	.0079	.0068	.0062	.0060	.0061	.0065	.0073	.0085
	0	.030	.018	.015	.013	.012	.012	.012	.013	.014	.015
	20	.049	.030	.024	.021	.020	.019	.019	.020	.022	.024
	40	.076	.046	.038	.034	.032	.032	.032	.032	.034	.036
	60	.122	.072	.056	.049	.046	.045	.045	.046	.047	.049
	80	.168	.098	.077	.066	.061	.058	.057	.056	.057	.058
	100	.238	.135	.104	.090	.081	.076	.073	.072	.072	.074
	120	.334	.187	.139	.128	.106	.099	.096	.094	.092	.093
n-PENTANE	−20	.013	.0079	.0062	.0053	.0049	.0047	.0048	.0051	.0056	.0065
	0	.021	.013	.011	.010	.0096	.0095	.0099	.011	.011	.012
	20	.037	.023	.020	.017	.016	.016	.016	.016	.017	.019
	40	.056	.035	.029	.026	.024	.025	.025	.026	.028	.029
	60	.088	.055	.044	.039	.037	.035	.035	.035	.036	.037
	80	.126	.075	.060	.052	.048	.046	.045	.043	.046	.046
	100	.182	.105	.081	.071	.064	.060	.058	.058	.058	.059
	120	.255	.145	.110	.094	.084	.079	.077	.075	.075	.076
HEXANE	−20	.0027	.0020	.0019	.0019	.0020	.0022	.0024	.0027	.0030	.0034
	0	.0056	.0037	.0034	.0034	.0036	.0039	.0041	.0046	.0052	.0058
	20	.011	.0068	.0054	.0047	.0044	.0042	.0045	.0048	.0054	.0062
	40	.016	.011	.0087	.0081	.0080	.0080	.0083	.0086	.0091	.0098
	60	.027	.017	.013	.012	.012	.012	.012	.012	.013	.013
	80	.041	.025	.020	.018	.017	.016	.016	.017	.017	.018
	100	.066	.038	.029	.026	.024	.023	.023	.023	.024	.025
	120	.096	.055	.042	.036	.033	.030	.030	.030	.031	.032

APPENDIX 5B (Continued)

COMPOSITE K VALUES FOR RESERVOIR FLIUDS (CONT'D)

Temperature, °F	Pressure, psia									
	100	200	300	400	500	600	700	800	900	1000
240°F, MBP										
–20	.00028	.00033	.00035	.00039	.00041	.00047	.00050	.00060	.00074	.00078
0	.00096	.0007	.00061	.00069	.00080	.00089	.0010	.0011	.0014	.0016
20	.0023	.0014	.0012	.0010	.00096	.00098	.0011	.0012	.0015	.0017
40	.0036	.0024	.0021	.0020	.0020	.0021	.0022	.0024	.0027	.0029
60	.0060	.0039	.0031	.0026	.0027	.0032	.0037	.0042	.0044	.0046
80	.010	.0061	.0052	.0046	.0045	.0048	.0050	.0053	.0054	.0056
100	.016	.010	.0085	.0074	.0072	.0074	.0076	.0080	.0086	.0088
120	.030	.017	.013	.012	.010	.010	.010	.011	.011	.012
260°F, MBP										
–20	.00018	.00022	.00024	.00027	.00029	.00034	.00035	.00044	.00054	.00058
0	.00064	.00047	.00043	.00049	.00056	.00064	.00072	.00084	.0010	.0012
20	.0016	.00096	.00086	.00072	.00068	.00070	.00078	.00086	.0011	.0013
40	.0025	.0017	.0015	.0015	.0015	.0015	.0016	.0018	.0021	.0021
60	.0043	.0028	.0022	.0019	.0020	.0024	.0028	.0031	.0033	.0036
80	.0072	.044	.0039	.0034	.0033	.0035	.0037	.0040	.0041	.0043
100	.012	.0076	.0063	.0056	.0055	.0056	.0058	.0061	.0066	.0069
120	.022	.013	.0010	.0088	.0079	.080	.0081	.0040	.0092	.0097
280°F, MBP										
–20	.00011	.00015	.00016	.00018	.00020	.00024	.00025	.00032	.00040	.00042
0	.00044	.00032	.00030	.00033	.00040	.00046	.00052	.00060	.00074	.00090
20	.0011	.00067	.00060	.00051	.00048	.00050	.00056	.00077	.00079	.00096
40	.0017	.0012	.0010	.0010	.0011	.0011	.0012	.0013	.0015	.0016
60	.0030	.0020	.0016	.0014	.0014	.0017	.0021	.0024	.0025	.0028
80	.0053	.0032	.0029	.0025	.0025	.0026	.0027	.0031	.0033	.0034
100	.0089	.0055	.0047	.0041	.0042	.0042	.0043	.0047	.0052	.0054
120	.0017	.0095	.0075	.0068	.0060	.0062	.0064	.0070	.0072	.0075
300°F, MBP										
–20	.000070	.000097	.00011	.00013	.00014	.00017	.00018	.00023	.00028	.00030
0	.00028	.00022	.00020	.00024	.00029	.00033	.00037	.00044	.00054	.00066
20	.00076	.00046	.00043	.00036	.00034	.00036	.00041	.00057	.00060	.00072
40	.0012	.00081	.00074	.00074	.00078	.00082	.00090	.00095	.0011	.0012
60	.0021	.0014	.0011	.00099	.0010	.0013	.0015	.0017	.0019	.0021
80	.0038	.0024	.0021	.0019	.0018	.0020	.0021	.0023	.0024	.0025
100	.0066	.0041	.0034	.0031	.0032	.0032	.0034	.0035	.0040	.0042
120	.013	.0072	.0057	.0051	.0045	.0047	.0049	.0054	.0056	.0060

6

WATER-HYDROCARBON PHASE BEHAVIOR

Water and hydrocarbons are natural companions. Hydrocarbons are formed in a water environment and are in equilibrium with water at reservoir conditions.

Liquid water and hydrocarbons are essentially immiscible in each other. Their slight mutual solubility is important for hydrate prediction but has no measurable effect on their vapor-liquid equilibrium behavior. At 25°C and atmospheric pressure the mol fraction of a paraffin like n-pentane in water is about 10^{-5}. Volume 3 discusses solubility parameters in significant detail.*(6.1)*

If both liquid water and hydrocarbons - and gas - are present in a system there will be two liquid phases. The amount of water vapor in the gas will be governed by gas composition and the vapor pressure of the liquid water phase.

Two calculations are primary to the water-hydrocarbon system: calculation of the water content of the gas and prediction of the conditions under which hydrates will form. Prevention of hydrates by dehydration or inhibition is important in petroleum operations.

WATER CONTENT OF GASES

The water content of a gas depends on system pressure and temperature and the composition of the water containing gas. The published record in this area goes back to the work of Poynting in 1881. Current methods of calculation include the use of:

1. Partial pressure and partial fugacity relationships.
2. Empirical plots of water content versus P and T.
3. Corrections to (2) for the presence of contaminants like hydrogen sulfide, carbon dioxide and nitrogen.
4. PVT equations of state.

Partial Pressure and Fugacity

If Raoult's Law (discussed in Chapter 5) is applied to water, it is written

$$P\,y_w = P_v\,x_w \tag{6.1}$$

Where: P = absolute system pressure
y_w = mol fraction water in the vapor phase
P_v = vapor pressure of water at system temperature
x_w = mol fraction water in the liquid water phase (= 1.0)

The liquid mol fraction can be taken as unity because of the immiscibility of the liquid phases. Thus, for a known pressure and water vapor pressure the mol fraction water in the vapor phase is found from Equation 6.1.

However, this equation is valid only at low pressure where the ideal gas law is valid. Equation 6.1 is recommended for system pressures up to about 400 kPa [60 psia].

Empirical Plots

Until the early 1950's most of the commercial natural gas had been processed in NGL units for its heavier components and most of its contaminants. For this *lean, sweet natural gas* the log of water content (W) was plotted versus P and T. It was found that a plot of log W versus 1/T approximated a straight line at a given pressure: most plots use log W versus T.

Figures 6.1 and 6.1(a) are this type of plot. They are a composite based on all earlier charts and other data available to the author.*(6.2-6.4)* The water content shown is the maximum gas can hold at the P and T shown. It is fully saturated; the relative humidity is 100% or, stated another way, the temperature is the water dewpoint temperature of the gas at the concentration and pressure shown.

The concentrations in mass per unit standard volume are related to mol fraction, y, as follows:

$$\text{kg water}/10^6\ \text{std m}^3 = (y_w)(18)(41\,740) = 751\,320\,y_w$$

$$\text{lbm water/MMscf} = (y_w)(18)(2636) = 47\,448\,y_w$$

Where: 18 = mol wt of water
41 740 = kmol gas/10^6 std m^3 (@ 15°C and 100 kPa)
2636 = lb-mol gas/MMscf (@ 60°F and 14.7 psia)

Comparison of the water content at various points in the system serves many purposes. One is to determine the water loading for dehydration. Another is to establish how much water has been condensed as liquid in the line.

Suppose at one point in a system the water content (from a correlation) is 200 kg/million std m^3. At some point downstream the same correlation gives a maximum water content of 100 kg/million std m^3. What is the conclusion? If no water has been withdrawn, there is 100 kg of liquid water in the line per million std m^3 of gas flowing. This water is available to form hydrates unless it is removed or inhibited. It also is a primary source of corrosion-erosion problems. The purpose of dehydration is to prevent such water condensation.

Figures 6.1 and 6.1(a) are satisfactory for most applications involving lean, sweet natural gas. The likely error of 6-10% probably is more accurate than the data the correlation is being applied toward. However, as the amounts of carbon dioxide and hydrogen sulfide increase the accuracy becomes poorer. It is good practice to make a correction for these contaminants even though it may be small when concentrations and pressures are low.

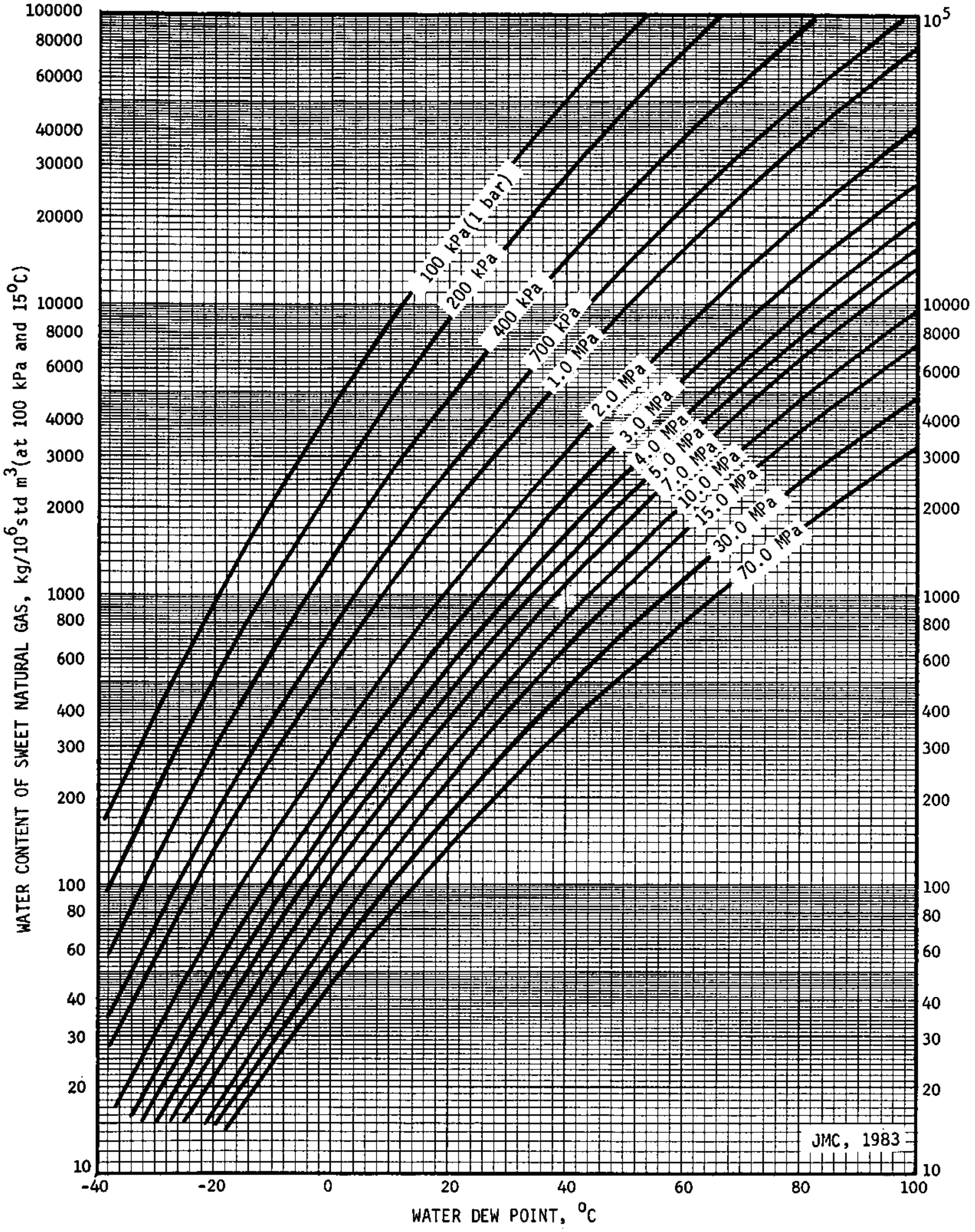


Figure 6.1 Water Content of Sweet, Lean Natural Gas

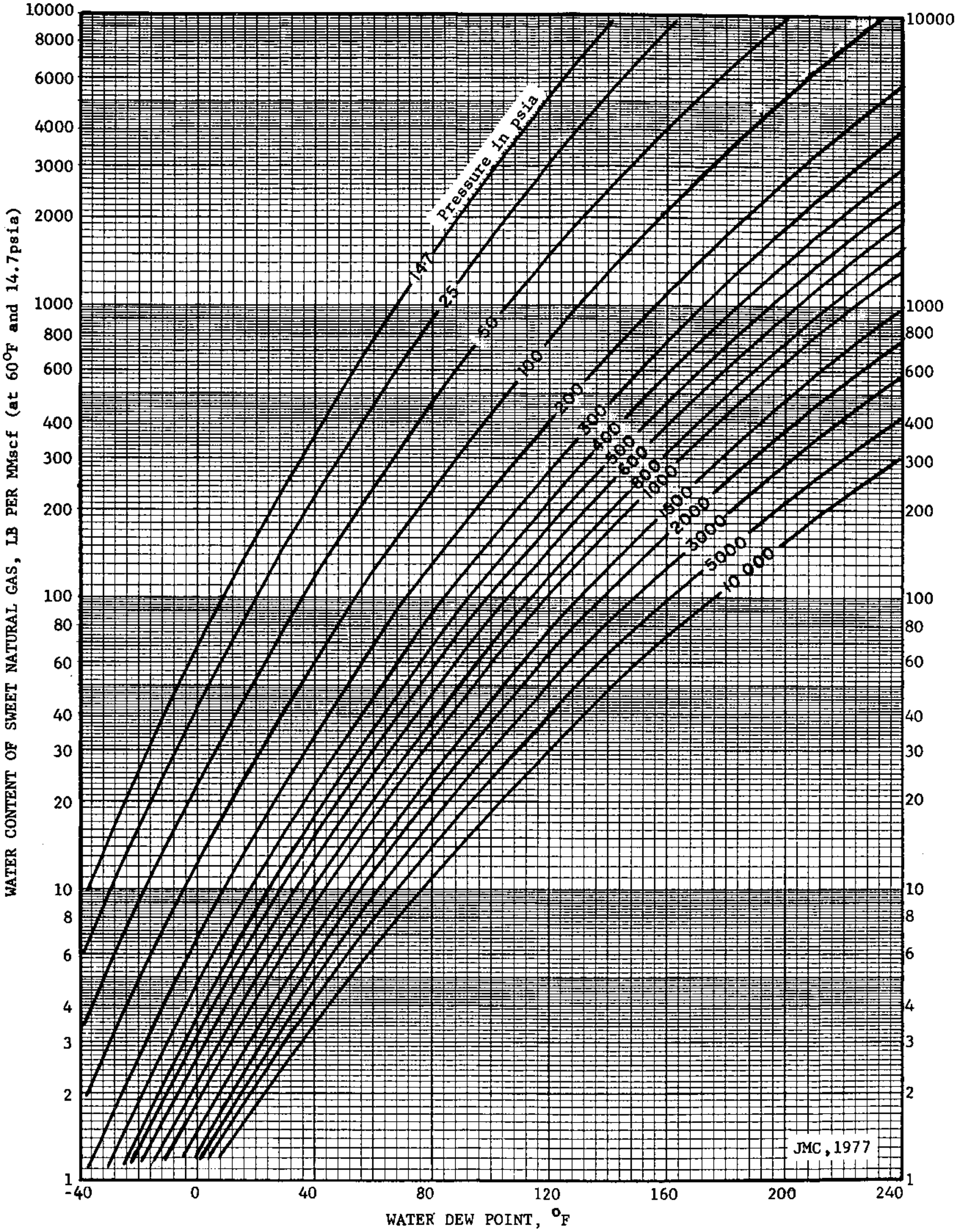


Figure 6.1(a) Water Content of Sweet, Lean Natural Gas

WATER CONTENT OF SOUR GASES

There are several available methods for determining the water content of gases containing H_2S and CO_2. Two of these use Equation 6.2 by multiplying the water content of the pure sour component by its mol fraction in the mixture. Figures 6.2 and 6.3 show what is called the "effective water content." These curves were based on the pure sour component data but were adjusted so that Equation 6.2 matched measured data on high pressure streams containing more than 20% H_2S.

As we have compared the results of this method with others we find that it is consistently high. It is the maximum possible value that can occur within a range of possible values.

$$W = y\,W_{hc} + y_1\,W_1 + y_2\,W_2 \tag{6.2}$$

Where: W = water content of gas
W_{hc} = water content of hydrocarbon part of gas from Figure 6.1
W_1 = water content of CO_2 from Figure 6.2 or 6.4
W_2 = water content of H_2S from Figure 6.2 or 6.5
$y = 1 - y_1 - y_2$,
y_1 = mol fraction of CO_2,
y_2 = mol fraction of H_2S

A second correlation using Equation 6.2 is based on the data of Sharma discussed on the following pages. Figures 6.4 and 6.5 were obtained by cross-plotting and smoothing Sharma's binary data for methane, CO_2, and H_2S.(6.5)

SRK Sour Gas Correlation

Figure 6.6 is another correlation for estimating sour gas water content.(6.6) The charts shown were calculated from the SRK equation of state assuming that the hydrocarbon portion of the gas was methane. It was assumed also that CO_2 had 75% of the water content of H_2S at the same conditions. One thus multiplies the percent CO_2 by 0.75 and adds the result to the percent H_2S to use the charts. The water content shown in API bbl/MMscf can be converted as follows:

$$\text{lbm/MMscf} = (350)(\text{bbl/MMscf})$$

$$\text{kg}/10^6 \text{ std m}^3 = (5543)(\text{bbl/MMscf})$$

Figure 6.6 is a quick way to estimate sour gas water content. As shown, however, it also is limited somewhat by its reliance on binary data.

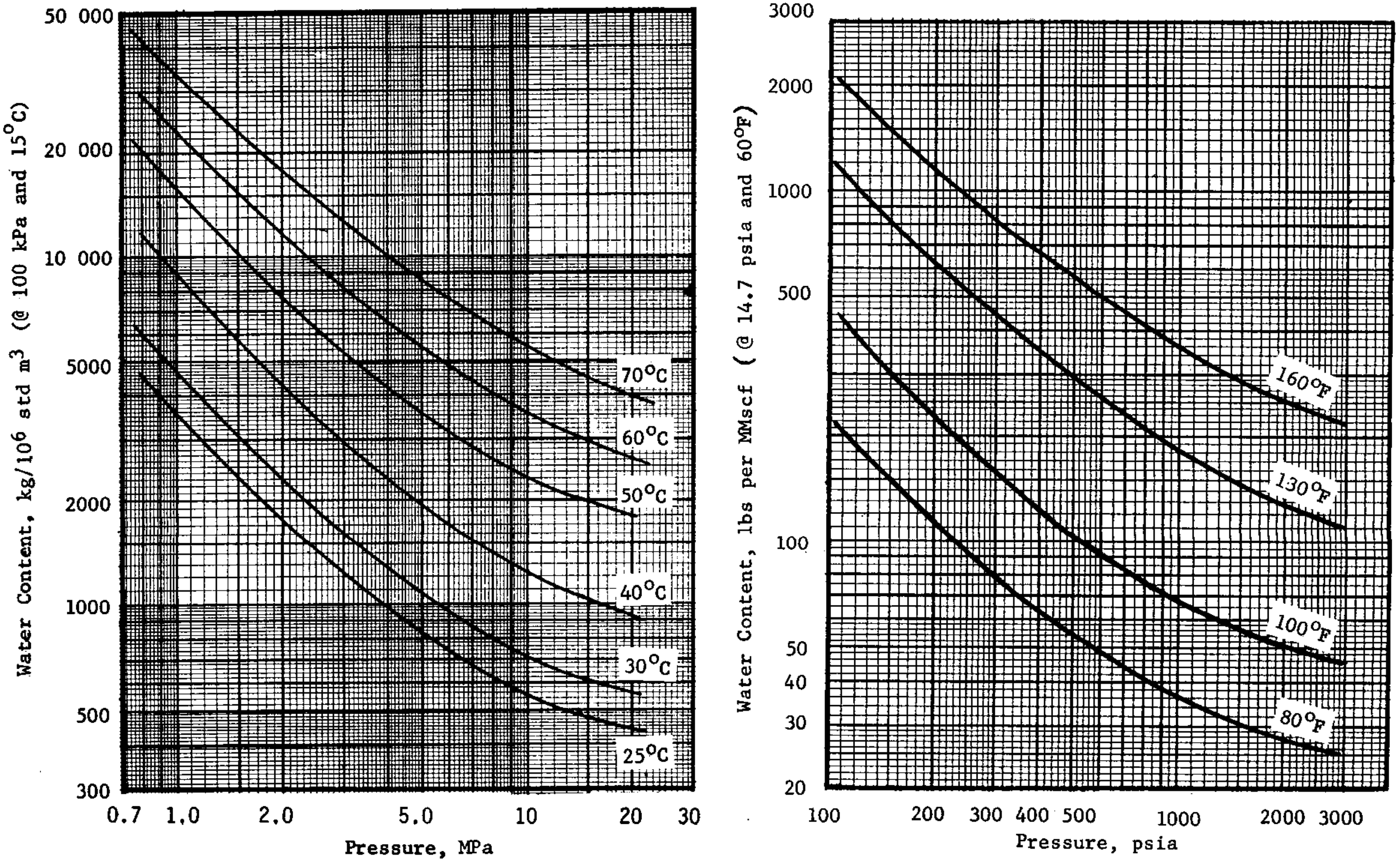


Figure 6.2 Effective Water Content of CO_2 in Saturated Natural Gas Mixtures

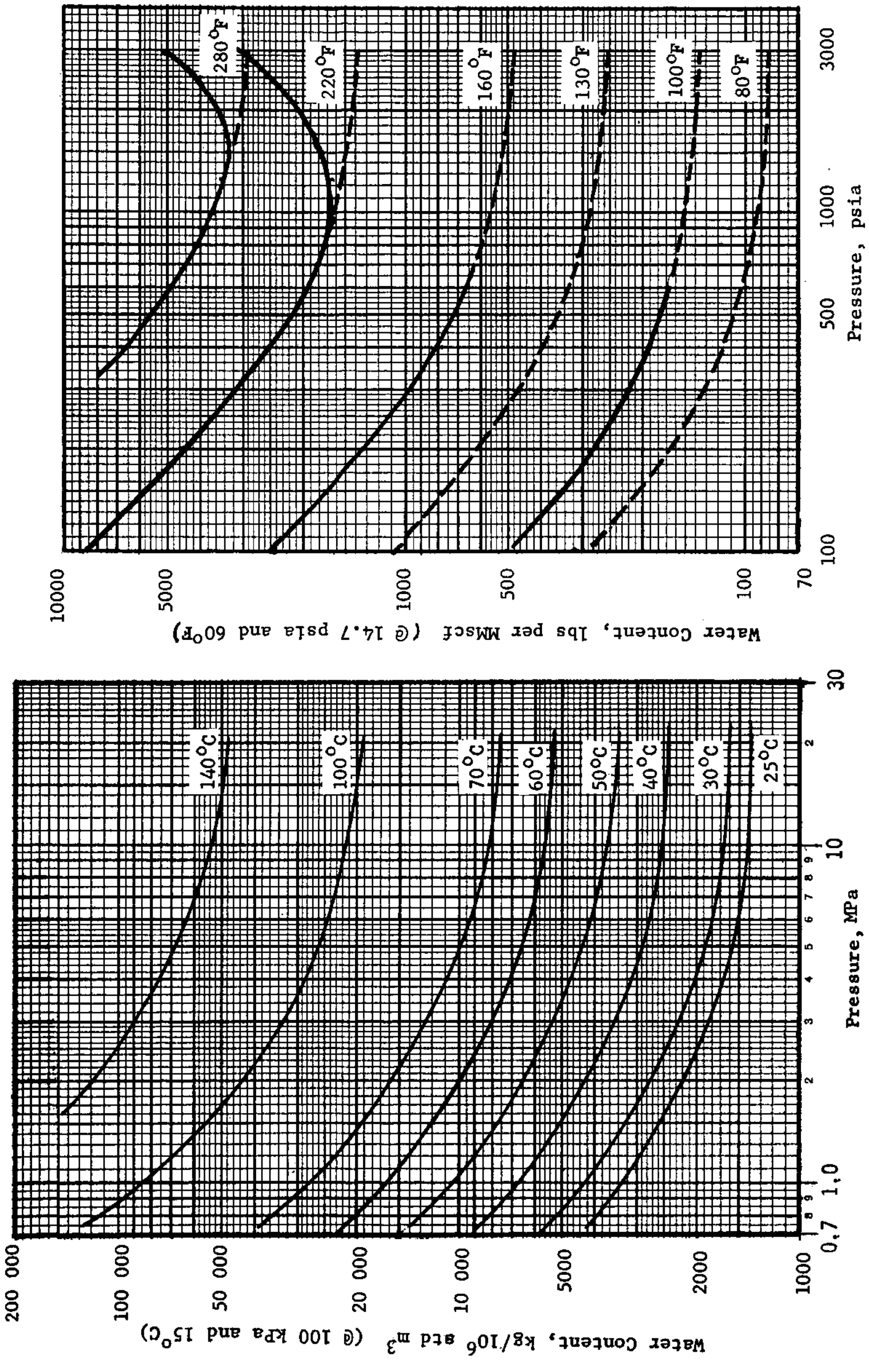


Figure 6.3 Effective Water Content of H_2S in Saturated Natural Gas Mixtures

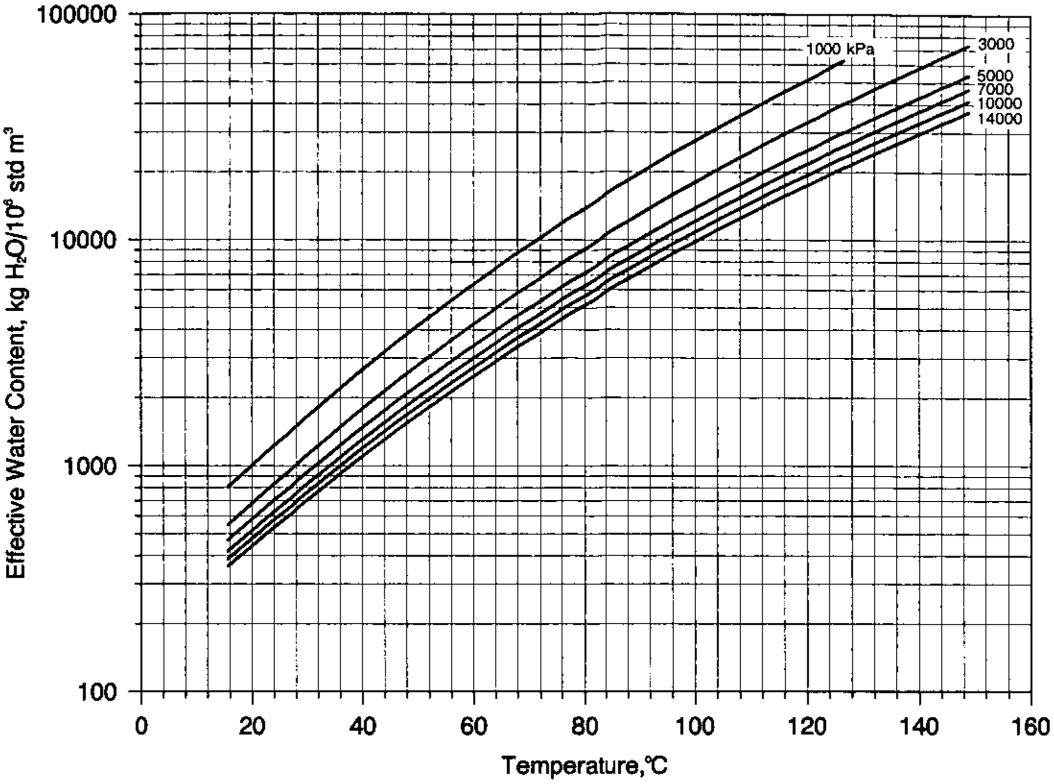


Figure 6.4 Water Content Contribution of CO_2

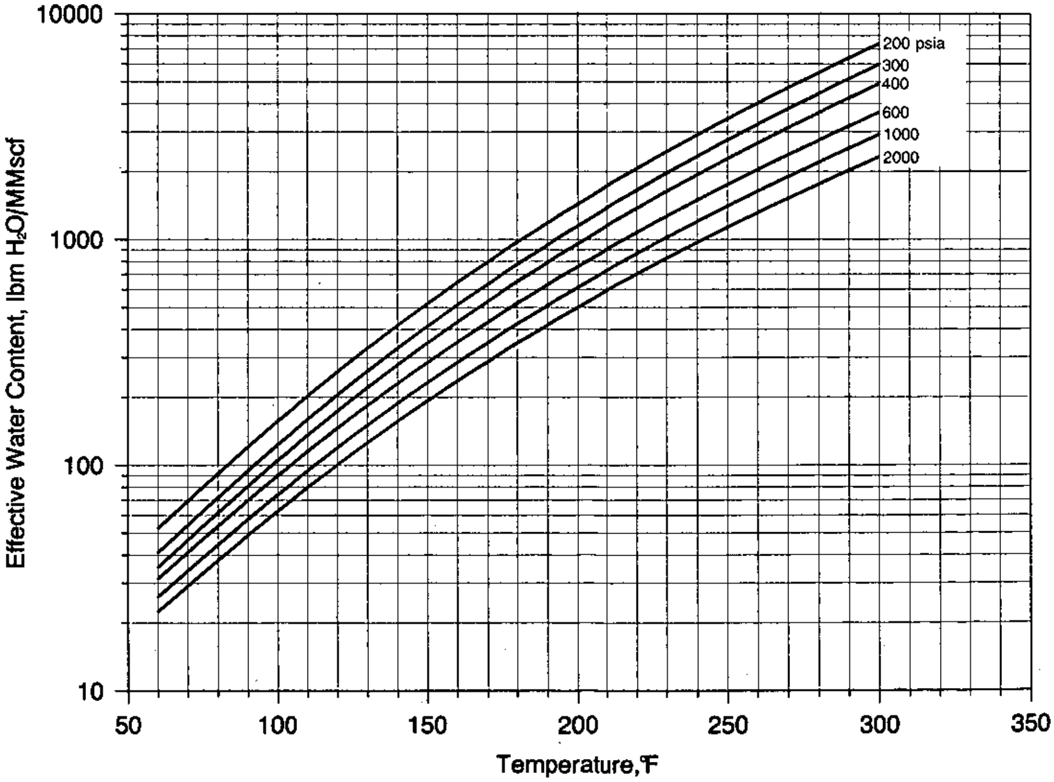


Figure 6.4(a) Water Content Contribution of CO_2

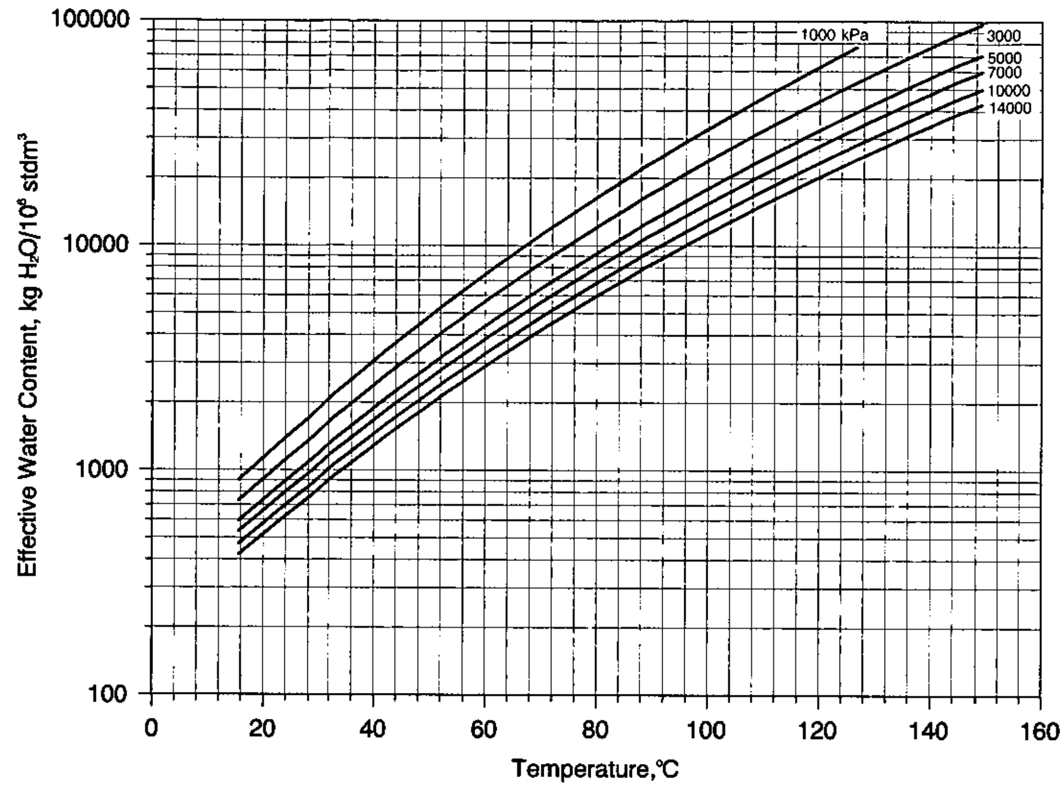


Figure 6.5 Water Content Contribution of H_2S

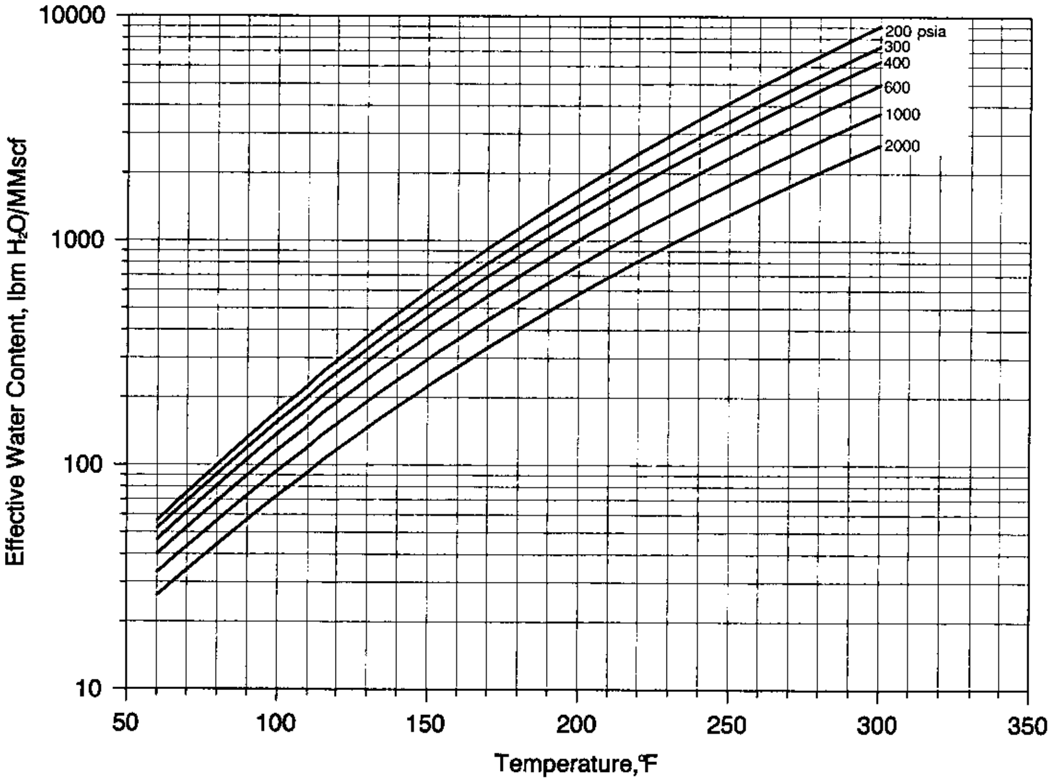


Figure 6.5(a) Water Content Contribution of H_2S

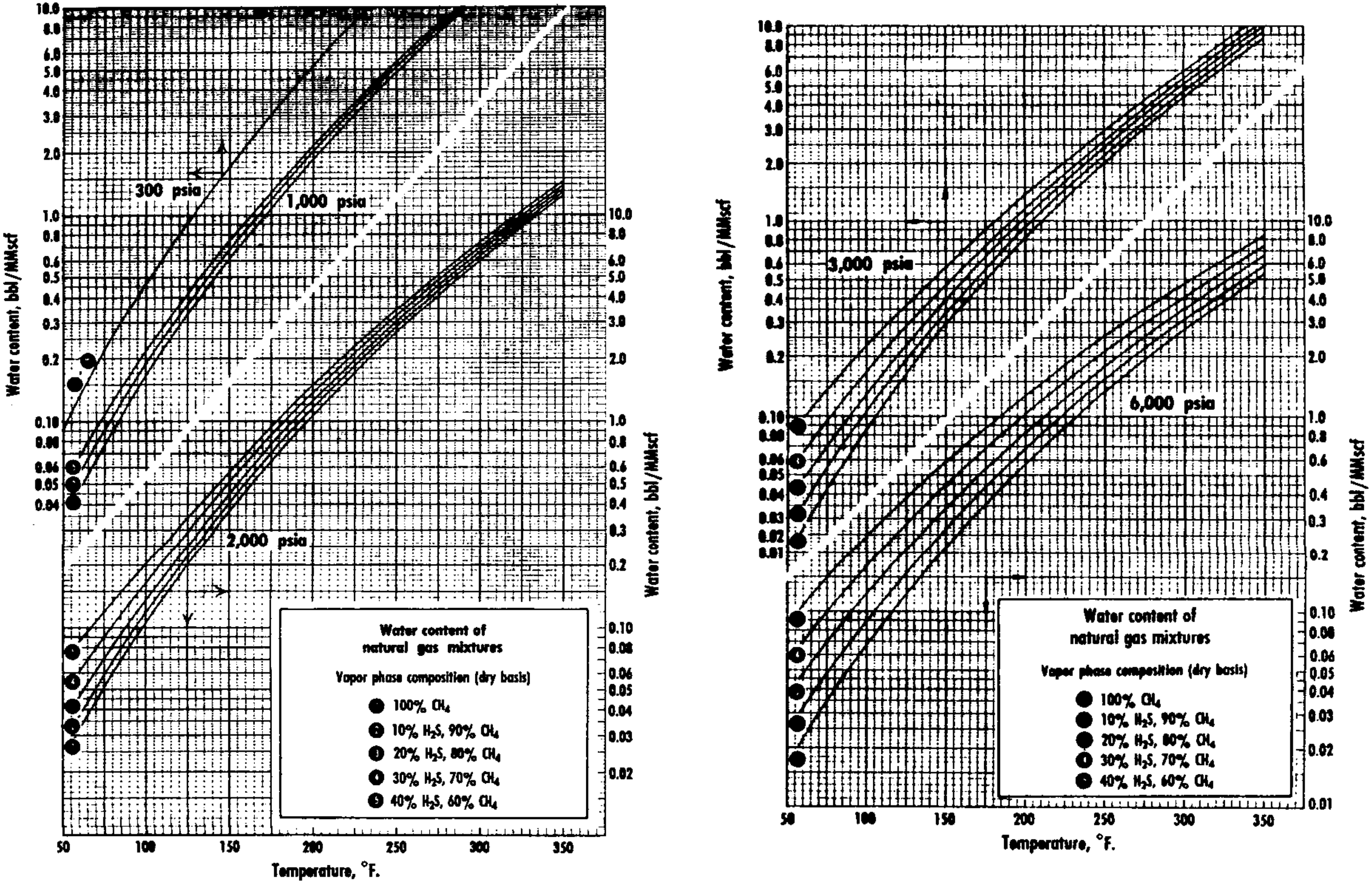


Figure 6.6 Another Correlation for Sour Natural Gases

Another Equation of State Approach

Suresh Sharma and the author have combined the Eykman Molecular Refraction concepts with standard physical chemistry equations to provide a basic correlation for sour gases. It too is empirical. It is more tedious manually than the others but represents a reliable alternative approach in the suite of methods presented. All of the figures shown herein are of course unnecessary when using the computer program for this method.

The first workable method for predicting water content was based on the concepts of EMR shown in Chapter 3 and standard physical chemistry principles.(6.7, 6.8) The method has been designed as a computer solution but can be performed manually using Figures 6.7-6.9

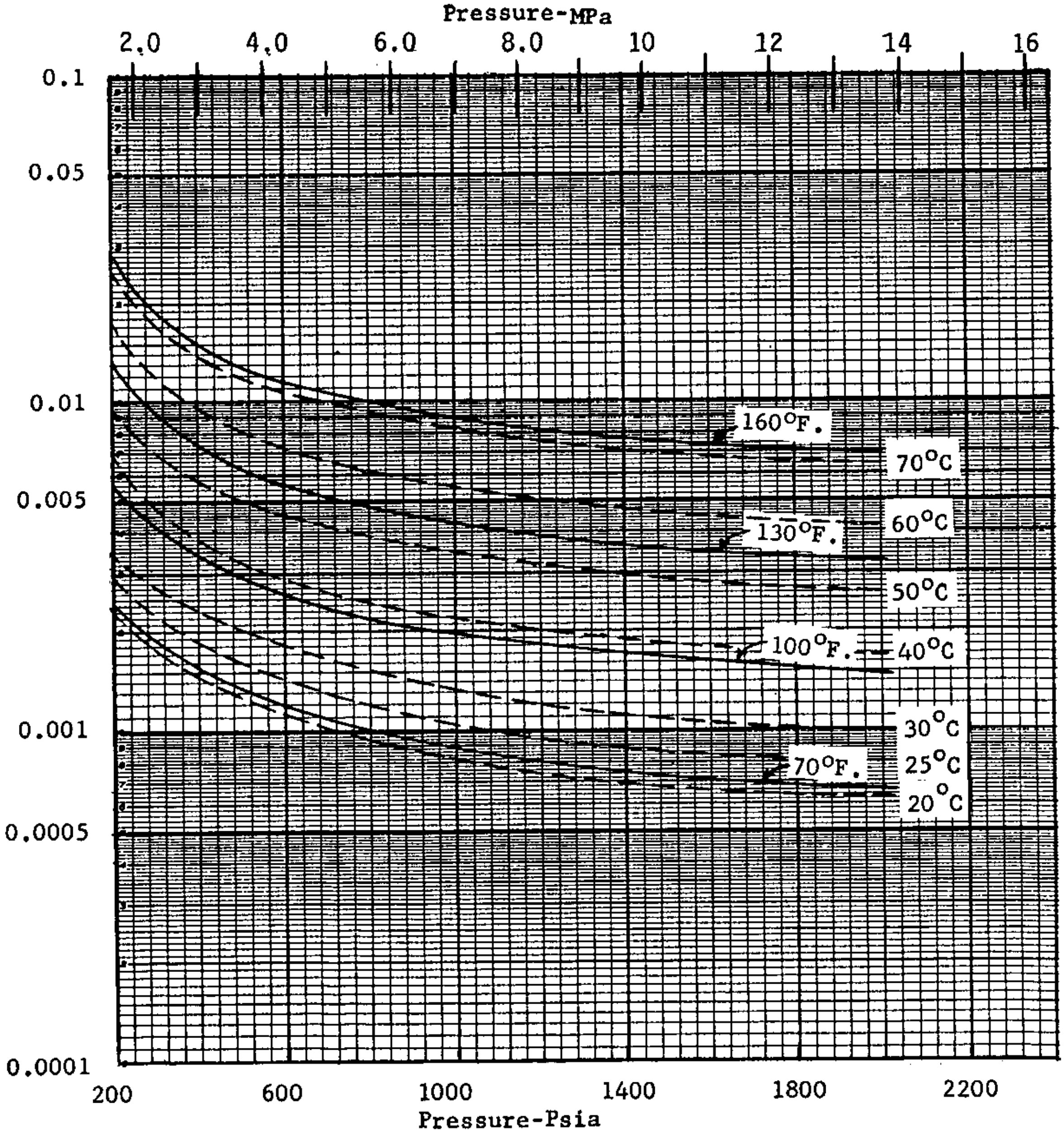


Figure 6.7 **Constant 'k" as a Function of Pressure and Temperature**

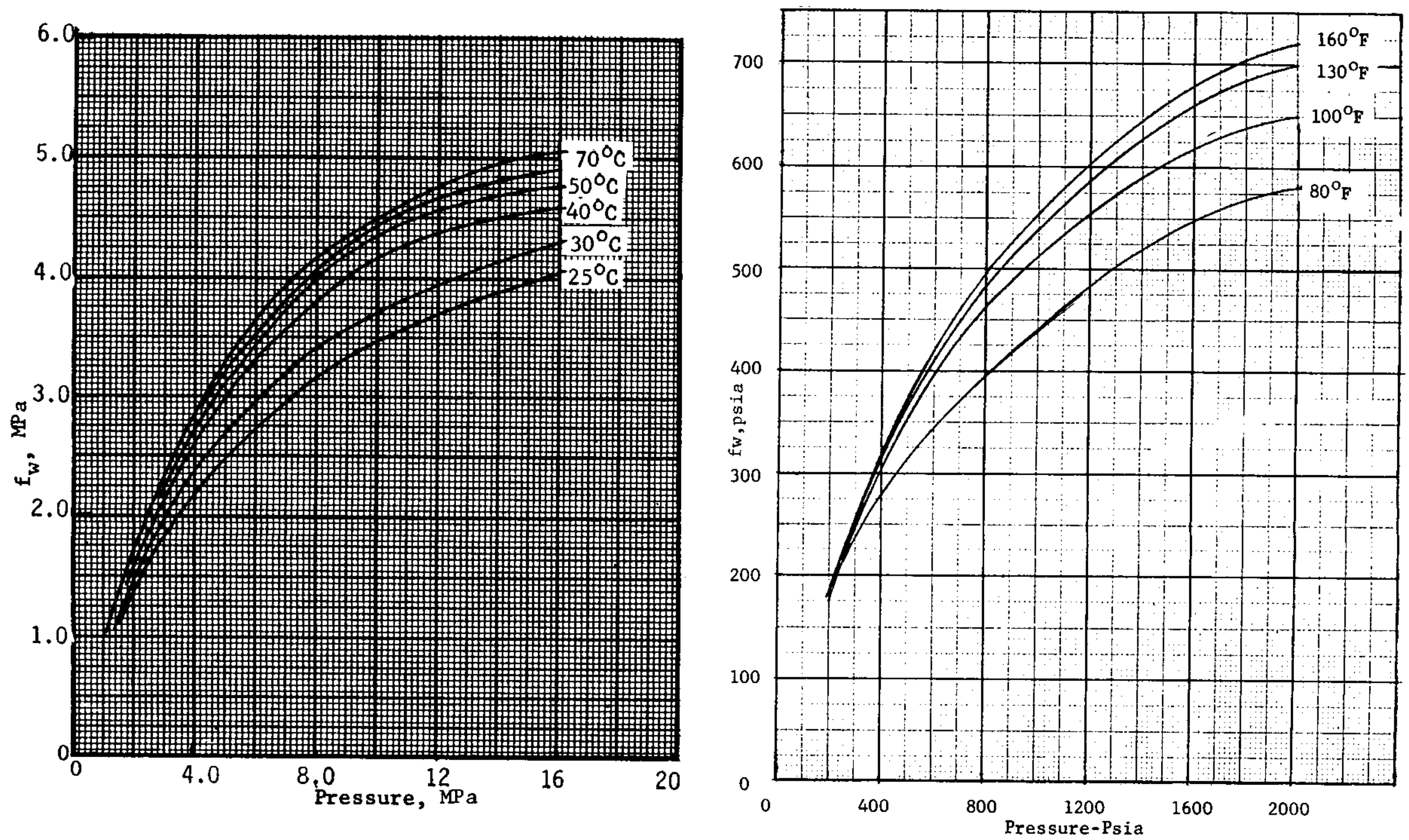


Figure 6.8 Water Fugacity as a Function of Pressure and Temperature

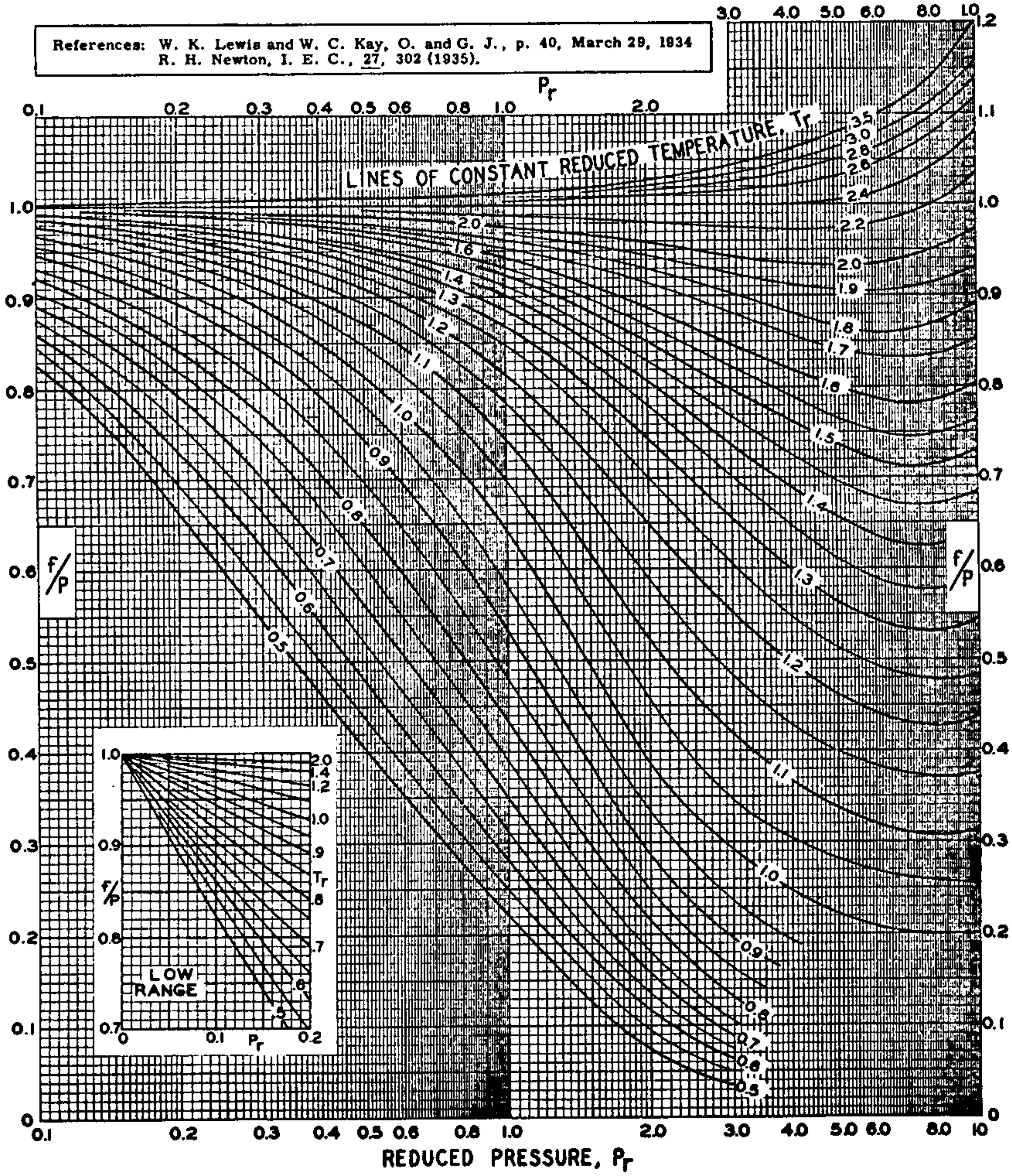


Figure 6.9 General Fugacity Coefficient Chart

The procedure is as follows.

1. Calculate the critical pressure and temperature of the mixture, using the technique of McLeod and Campbell summarized in Chapter 3, which utilizes molecular refraction as a third parameter.
2. Calculate reduced pressure and temperature for the gas system. Determine "Z" from EMR approach.
3. Find the value for constant "k" from Figure 6.7, or calculate by the equation.

$$k = \left(\frac{P_w^o}{P}\right)\left(\frac{f_w^o/P_w^o}{f_w/P}\right)\left(\frac{P}{P_w^o}\right)^{0.0049} \tag{6.3}$$

Where: P_w^o = vapor pressure of water at system temperature (T)
P = system pressure
f_w^o = fugacity of water at P_w^o and T
f_w = fugacity of water at P and T
0.0049 = semi-empirical constant

4. Determine the fugacity of water (f_w) from Figure 6.8, or calculate from Figure 6.9, using P_c and T_c for water.
5. Determine value of (f/P), the fugacity coefficient from Figure 6.9, using the reduced parameters from Step 2. Calculate "f."
6. Determine the water content by the equation

$$y = k\,(f_w/f)^Z \tag{6.4}$$

Where: y = mol fraction of water in vapor

In solving Equation 6.4 note that the Z must be from the EMR combination rule and no other source.

I use all of the above methods when calculating a sour gas water content. The reason – the random behavior of such systems. The repeatability of data is very poor, even in the laboratory. One should recognize this in applying such information.

You likely will get four different answers with the four correlations above. If dehydrator design or operation is involved, I suggest you use the highest of the results to be on the safe side.

Effect of Nitrogen and Heavy Ends

Nitrogen will hold less water than methane. At pressures to about 7.0 MPa [1000 psia] the water content of nitrogen is about 6-9% less than for methane. This deviation increases with pressure. Thus, including nitrogen as a hydrocarbon is practical and offers a small safety factor.

The presence of heavy ends tends to increase the water capacity of the gas. Once again the deviation is relatively small at normal system pressures. The nitrogen and heavy end effects tend to cancel out each other in many production systems.

Example 6.1: Calculate the saturated water content of a gas with the analysis shown below at 1100 psia and 120°F. EMR Z = 0.79, P_c = 867 psia, T_c = 436°R.

Comp.	y_i
N_2	0.0046
CO_2	0.0030
H_2S	0.1438
C_1	0.8414
C_2	0.0059
C_3	0.0008
iC_4	0.0003
nC_4	0.0002
	1.0000

1. From Figure 6.1(a), W = 97 lbm/MMscf
2. $y_1 = 0.0030$, $y_2 = 0.1438$
   From Figure 6.1(a), W_{hc} = 97 lbm/MMscf
   From Figure 6.2, $W_1 = 130$
   From Figure 6.3, $W_2 = 230$
   W = (0.8532)(97) + (0.003)(130) + (0.1438)(230) = 116 lbm/MMscf
3. From Figure 6.4(a), $W_1 = 120$
   From Figure 6.5(a), $W_2 = 150$
   W = (0.8532)(97) + (0.003)(120) + (0.1438)(150) = 105 lbm/MMscf
4. Effective % H_2S = (0.3)(0.75) + (14.38) = 14.6%
   From Figure 6.6, W = (350)(0.31) = 109 lbm/MMscf
5. From Figure 6.7, k = 0.0031
   From Figure 6.8, $f_w = 555$
   $P_r' = 1100/867 = 1.27$, $T_r' = 580/436 = 1.33$
   $f/P = 0.85$, $f = (0.85)(1100) = 935$
   $y = (0.0031)(555/935)^{0.79} = 0.00205$
   W = (0.00205)(47 448) = 97.3 lbm/MMscf

In Example 6.1 the water content from Equation 6.2 is greater than from Figure 6.1. This is expected for the sour gas in question. From my experience, a value as high as 116 is not likely but it does happen. The value of 97 represents a practical minimum.

Using probability principles, the numbers would be regarded as a distribution. The mode is the most likely value but any value within the range shown has some probability of occurring. This fact must be accepted in making any decision resulting from this calculation. Merely averaging the numbers or arbitrarily picking one as "sacred" is not consistent with the true facts of the matter.

One can estimate the water content, or relative saturation of the gas entering a dehydrator, from the water content correlations.

Example 6.2: Water is separated in an inlet scrubber to a compressor plant at 7.0 MPa and 35°C. The gas is then compressed and aftercooled to 14.0 MPa and 50°C before entering a dehydrator. Is the gas saturated?

From Figure 6.1, water content at 7.0 MPa and 35°C = 810 kg/10^6 std m^3

From Figure 6.1, water content at 14.0 MPa and 50°C = 1050 kg/10^6 std m^3

The gas is capable of holding 1050 kg but has only 810 kg. Therefore, it is not saturated. There is less than the maximum amount of water to be removed. Also, there is less driving force available to aid water removal. One must consider both factors.

In other cases the maximum water capacity of the gas downstream may be less than that upstream. In this case water must condense in the section involved. The water content correlation may be used to estimate the amount of liquid water present.

If no separation occurs between the wellhead and the dehydration plant, it is reasonable to assume that the gas was saturated at the formation pressure and temperature.

Saturated Water Content in Equilibrium with Hydrates

Figure 6.1 is based on the assumption that the condensed water phase is a liquid. However, at temperatures below the hydrate temperature of the gas, the "condensed" phase will be a solid (hydrate). The water content of a gas in equilibrium with a hydrate will be lower than equilibrium with a metastable liquid.

Hydrate formation is a time dependent process. The rate at which hydrate crystals form depends upon several factors including gas composition, presence of crystal nucleation sites in the liquid phase, degree of agitation, etc. During this transient "hydrate formation period" the liquid water present is termed "metastable liquid." Metastable water is liquid water which, at equilibrium, will exist as a hydrate.

References 6.9-6.11 present experimental data showing equilibrium water contents of gases above hydrates. Data from Reference 6.10 is presented in Figure 6.10. For comparative purposes, the "metastable" water content of a sweet gas from Figure 6.1 is also shown. The water content of gases in the hydrate region is a strong function of composition. Figure 6.10 should not be extrapolated to other compositions.

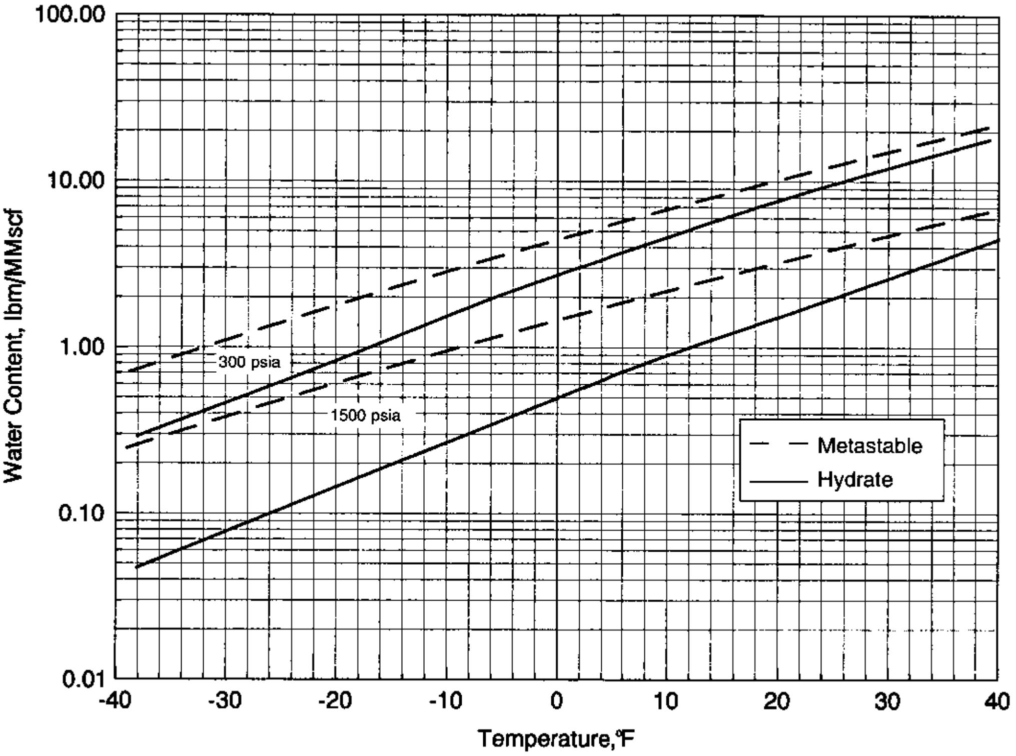


Figure 6.10 Water Content of 5.31% C_3 - 94.69% C_1 Gas in Equilibrium with Hydrate

When designing dehydration systems, particularly TEG systems to meet extremely low water dewpoint specifications, it is necessary to determine the water content of the gas in equilibrium with a hydrate using a correlation like that presented in Figure 6.10. If a metastable correlation is used, one will overestimate the saturated water content of the gas at the dewpoint specification. This, in turn, may result in a dehydration design which is unable to meet the required water removal. Where experimental data is unavailable, utilization of a sound thermodynamic-based correlation can provide an estimate of water content in equilibrium with hydrates.

WATER MONITORS

Measurement of the water content of gas is a normal part of the gas conditioning process. There are many instruments on the market for this purpose, in a wide price range. The choice is not routine. Some are difficult, or certainly inconvenient, to calibrate. Some will not stand hard physical treatment. Others are not suitable for hazardous atmospheres.

The grandfather of this family of instruments is what is often called the Bureau of Mines Dewpoint Tester. It consists of a polished mirror that can be cooled at a steady rate. When the first drops of liquid "dew" appear visually on this mirror the technician records the temperature and pressure. It is simple and rugged. But, it depends on the visual judgment of the technician. Different people get different results.

One problem is the fact that hydrocarbons often condense at a higher temperature than water and flood the mirror. That little opaque spot may be hard to detect. One solution for this is to place an activated carbon chamber in the test line ahead of the tester. Carbon has little affinity for water. Once it has been saturated the gas leaving will be representative of the gas being tested. At best though this type of tester is not a completely satisfactory instrument when used as a guide in custody transfer operations.

There is also a visual type based on expanding gas across a valve to produce a fog. One finds the vanishing point of the fog to determine the dewpoint. It is a low pressure instrument requiring gas expansion.

A more sophisticated form of the mirror concept uses a gold plated mirror bonded to a copper thermocouple holder. As this assembly is cooled an optical sensing bridge detects the change in light level that occurs when dew forms on the mirror.

One of the better (and more expensive) instruments uses two hygroscopically coated quartz crystal oscillators which vibrate at nine million cycles per second. Water vapor is alternately adsorbed and desorbed on each crystal. The mass change involved is reflected by changes in frequency, which are compared electronically. Each crystal is exposed alternately to the moist sample gas for 30 seconds while the other is exposed to a dry reference gas.

Still another instrument uses an anodized aluminum strip providing an aluminum oxide layer on which a thin layer of gold has been deposited. Two conductors are used. When water adsorbs on the oxide surface it changes the dielectric properties. The impedance is measured and converted to water content in equilibrium with the sensor.

There are many variations of the electrolytic moisture analyzer available. The exact system varies with the manufacturer. One type consists of a glass tube with two platinum wires wound in a double helix on the inside surface. The space on this surface, between the wires, is coated with phosphorous pentoxide (P_2O_5). When moisture wets the P_2O_5 an electrolysis current is produced as a potential is applied to the wires. The magnitude of this current is used to measure water content.

Another instrument measures water content by measurement of the heat energy exchanged when gas is adsorbed or desorbed on the surface of a solid adsorbent. The sample is split into two streams, one of which is thoroughly dried.

One also can measure water content in a chromatograph. In one method the gas is passed through calcium carbide. The water reacts producing equivalent amounts of acetylene which can be measured in the chromatograph. Another chromatographic method measures the water content using two identical columns coated with glycerol.

Many of the instruments described briefly above may be satisfactory in a given case. None are perfect. Some require calibration which at best is a nuisance. If you have sour gases the choice becomes very limited and none may be truly satisfactory. The electrolytic cells are particularly susceptible to poisoning.

WATER CONTENT APPLICATIONS

Water content correlations are used primarily for dehydration calculations and to determine how much water (if any) will condense from the gas. The latter involves considerations of disposal, corrosion/erosion and hydrate inhibition.

The amount of condensed water is relative small. The real danger is underestimating the quantity. Be sure that your estimates are on the "safe" side of the possible range of values. The additional capital expenditure normally will be trivial. There is a tendency to predict flowing temperatures lower than they turn out to be. Notice in Figure 6.1 that the water content is a log scale. A change in temperature of only 10% in the ambient range causes about a 33% increase in water content. A common cause of poor dehydrator performance is underprediction of the water loading.

References 6.1-6.12 all address water content correlations; 6.12 is an excellent review of water content correlations.

GAS HYDRATES

Gas hydrates are a form of clathrate first discovered by Davy in 1810. Villard in 1888 presented performance data on hydrocarbon hydrates. Thus, the scientific study of these physical compounds has been very extensive.

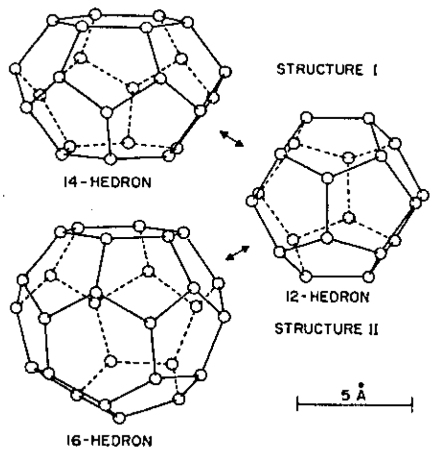


Figure 6.11 Schematic of Natural Gas Hydrate Lattice

A clathrate is any compound wherein guest molecules are entrapped in a cage structure composed of host molecules. With the natural gas hydrate the lattice is formed by water molecules as shown in Figure 6.11. The circles represent water molecules and the lines hydrogen bonds. Contained within each lattice is a series of cavities or cages that must be occupied by enough guest molecules to stabilize this lattice crystal.

X-ray diffraction analysis shows the gas hydrate crystal structure to be a derivative of the pentagonal dodecahedron, a twelve sided structure whose faces have five edges. Since the bonding angle is 108° and that of ice is 109.5°, the structure was long thought to be the probable basic hydrate building structure; yet, no orderly packing arrangement can be made with the regular pentagonal dodecahedron. Claussen proposed two separate crystal lattices of modified dodecahedron configurations designated as Structures I and II.

The basic lattice of Structure I has a cell constant of 12 A° and contains forty-six water molecules arranged to form eight voids, two of which are pentagonal dodecahedra. The remaining six voids are somewhat larger cavities having fourteen faces. Designated as tetrakaidecahedron, each large void has twelve pentagonal faces and two hexagonal faces. Molecules of methane are able to enter either size void, while a larger hydrocarbon such as ethane can be entrapped only in the larger voids. During formation it is also possible that some cavities are left unoccupied due to the lattice forming and closing when no hydrocarbon molecule is properly oriented. Other hydrocarbon molecules such as propane and larger are not entrapped in this structure due to their unfavorable size.

A second crystal form, Structure II, designated as a pentagonal dodecahedral-diamond lattice, is formed in the presence of hydrocarbons larger than ethane. For this structure, a unit cell having 136 water molecules and a cell constant of 17 A° is present. There are sixteen small pentagonal dodecahedral voids and eight larger voids. The large voids, termed hexakaidecahedron, have four hexagonal and twelve pentagonal faces. Entrapment of propane, n-butane, and isobutane occurs in these structures as well as possible entrapment of ethane and methane. It can be surmised that the absence of propane and heavier molecules would cause only Type I hydrates to be formed, while the presence of both ethane and propane would result in both. This is possible since the availability of excess water and hydrocarbons during hydrate formation may be limited.

It is apparent from the above that molecular size is an important factor. A second important factor is gas solubility. Solubility primarily affects the rate of clathration since it governs the statistical probability that a given molecule will be present when the lattice is closing. Hence, the greater the solubility of a given component, the faster a hydrate will form.

It is believed that the guest molecule dictates which type of structure will be formed, and its solubility governs the rate of formation. Size also affects the formation rate. This may be seen from the hydrate behavior of methane, hydrogen sulfide, and propane. At a given temperature the pure methane hydrate requires a higher pressure to form than the propane hydrate, although its solubility in water is considerably greater than that of propane. It is much easier for a small molecule to avoid entrapment as the cage is closing than a larger one. There are more methane molecules present, but their more active, random movement, and smaller size make clathration more difficult. The similarly sized hydrogen sulfide molecule is far more soluble than methane, forms in the same structure, and requires about one-twentieth the time to hydrate.

The actual imprisonment conditions are considered satisfied when the gas molecule is properly oriented within the water structure such that the aggregation of water and gas molecules have best utilized the available space. This corresponds to the state of lowest potential energy for the aggregation such that attractive, repulsive, rotational, and bonding forces are as much in equilibrium as possible. Once formed, the hydrogen bonded structure will not be broken until these forces become so unbalanced that the strength of the hydrogen bond is exceeded and fracturing occurs.

The guest molecules provide stability to the lattice structure in the same manner that liquid in the pores of a subsurface sand prevents subsidence. Not all cages must be full. Therefore, there is no specific formula for hydrates.

The following is representative of typical hydrate crystals formed by a given component.(6.13)

$CH_4\ 5.9\ H_2O$, $C_2H_6\ 8.2\ H_2O$, $H_2S\ 6.1\ H_2O$, $CO_2\ 6.0\ H_2O$

$N_2\ 6.0\ H_2O$, $C_3H_8\ 17.0\ H_2O$, $i\text{–}C_4H_{10}\ 17.0\ H_2O$

Normal butane can form a hydrate but it is very unstable. All paraffin hydrocarbons larger than the butanes are nonhydrate formers.

A natural gas mixture forms hydrates more readily than any pure component. It is necessary to have all three sizes of cages relatively full in order to stabilize the lattice. The cell constant quoted above is the length of one side of a cube which would fit into the cavities. So a mixture of molecular sizes is conducive to hydrate formation.

The individual hydrate crystals form in the liquid water phase. Only in the liquid phase does one have the necessary concentration and energy level to allow hydrogen atom bonding. These crystals are very small. But, they can grow by sharing the faces composed of 5- and 6-membered water molecules. It is only when this growth continues to form a plug are hydrates a known problem.

HYDRATE EQUILIBRIUM

Figure 6.12$^{(6.14)}$ shows the hydrate forming characteristics of common natural gas components. The lines shown are the hydrate forming conditions. Below and to the right of the line no hydrates can form. At some temperature the hydrate forming line becomes vertical. This happens at the point where the hydrate forming curve intersects the vapor pressure curve of that component.

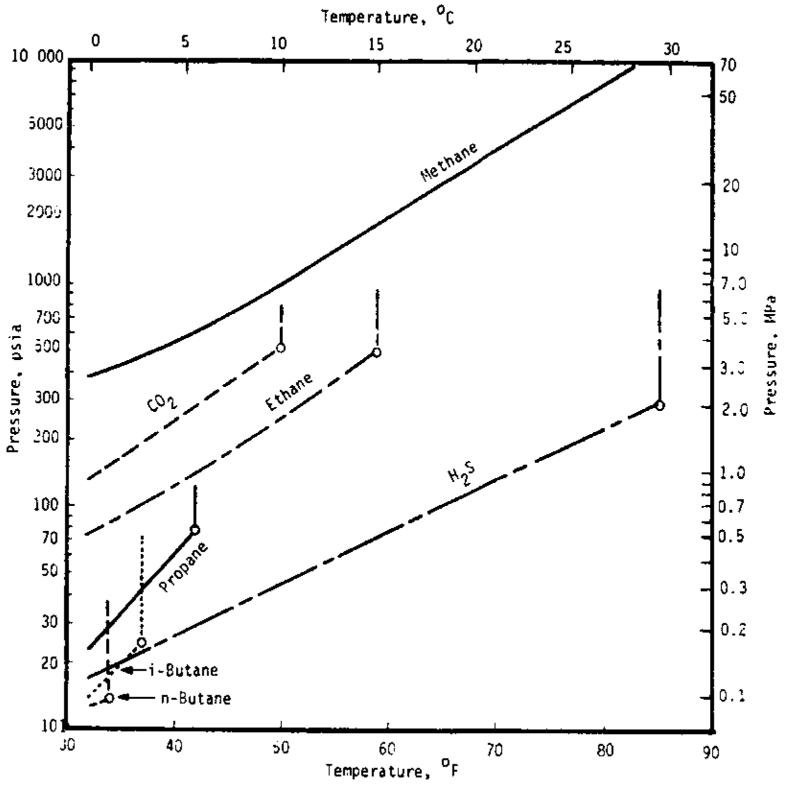
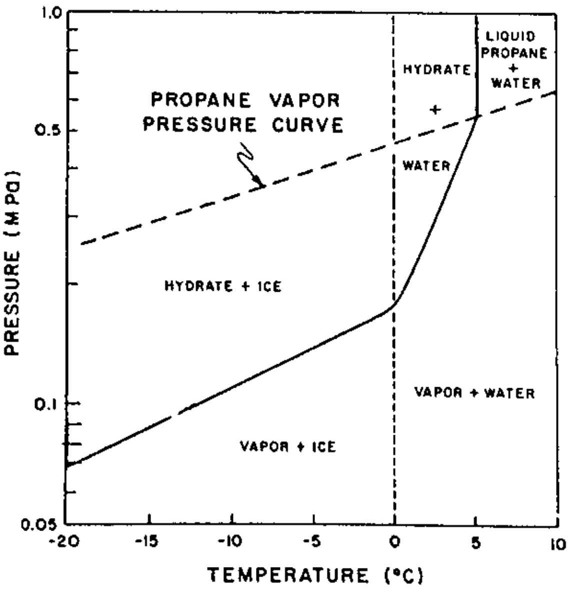


Figure 6.12 Hydrate Forming Conditions for Natural Gas Components



The figure at left illustrates the general behavior for propane.

Below 0°C [32°F] both ice and hydrate are present above the curve; only vapor and ice are below. Above the freezing point, hydrate is the only solid phase to the left of the hydrate forming curve. At the intersection of the vapor pressure and hydrate curves the latter becomes essentially vertical. This intersection establishes the *maximum hydrate forming temperature* for pure components.

The same general behavior occurs for a gas mixture, as shown in Figure 6.13. Line FEGC is the dewpoint line for the mixture. The slope of the hydrate curve (FGH) becomes increasingly vertical along the quadruple line (FG). The slope of the quadruple line is dependent upon the amount of hydrocarbon liquid in the system. Increasing amounts of liquid steepen

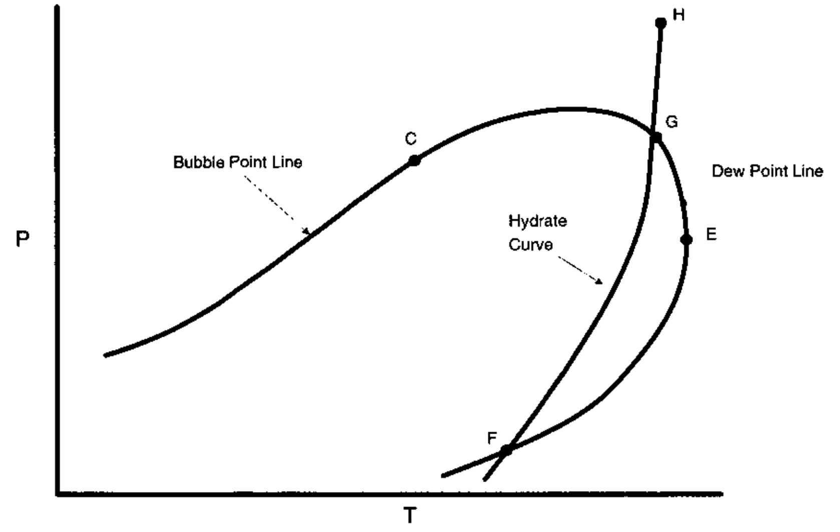


Figure 6.13 General Hydrate Formation Characteristics of a Gas Mixture

the slope. If the quadruple line intersects the phase envelope to the left of the critical point (bubble point line) the hydrate curve will be essentially vertical in the single phase region.

The general procedure for hydrate prediction involves two basic steps.

1. Use a suitable correlation to establish the position of the dewpoint line FEGC.
2. Use a suitable correlation to find hydrate line FGH.

The first step was covered in Chapter 5. The correlations which follow may be used for Step 2.

HYDRATE PREDICTION CORRELATIONS

All basic hydrate prediction correlations are for a system containing only gas and water in a static test cell that was rocked only to provide good equilibrium. In most cases the data shown are the hydrate melting conditions, a more reproducible data point than the formation point, and more reliable for calculation use.

Early correlations showed hydrate forming curves as a function of gas relative density. Although they are simple to use and may happen to apply for a lean gas, they are not recommended for engineering calculations.

Vapor-Solid Equilibrium Constants

Katz and his co-workers developed a set of $K_{v\text{-}s}$ values for hydrate prediction.(6.15-6.18) Figures 6.14- - 6.18 are the $K_{v\text{-}s}$ charts for use of this method as redrafted by the GPSA and modified for use herein. Figure 6.17 replaces the figure in the original work.(6.19)

In spite of some theoretical flaws in the method it has proven very reliable up to about 7 MPa [1000 psia]. Some of the curves do not exceed this pressure. With others the accuracy begins to decrease as the values converge to $K_{v\text{-}s} = 1.0$.

In the original work it was assumed that nitrogen was a nonhydrate former and that n-butane had the same $K_{v\text{-}s}$ value as ethane. Later work has shown these to be technically incorrect. A $K_{v\text{-}s}$ curve is now provided for n-butane, however, as a practical matter using $K_{v\text{-}s}$ = infinity for nitrogen gives very satisfactory results.

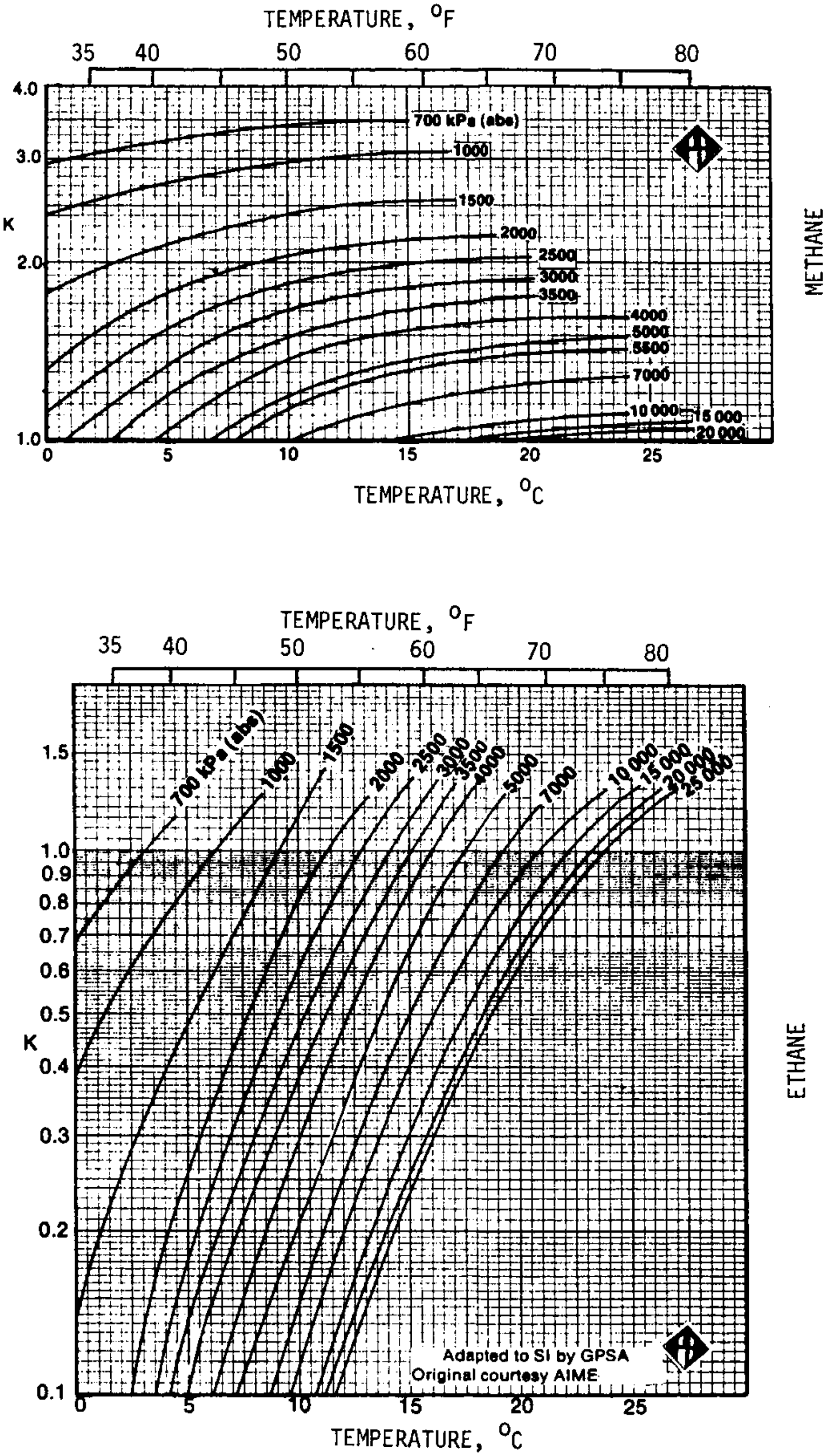


Figure 6.14 Vapor-Solid K Values for Methane and Ethane

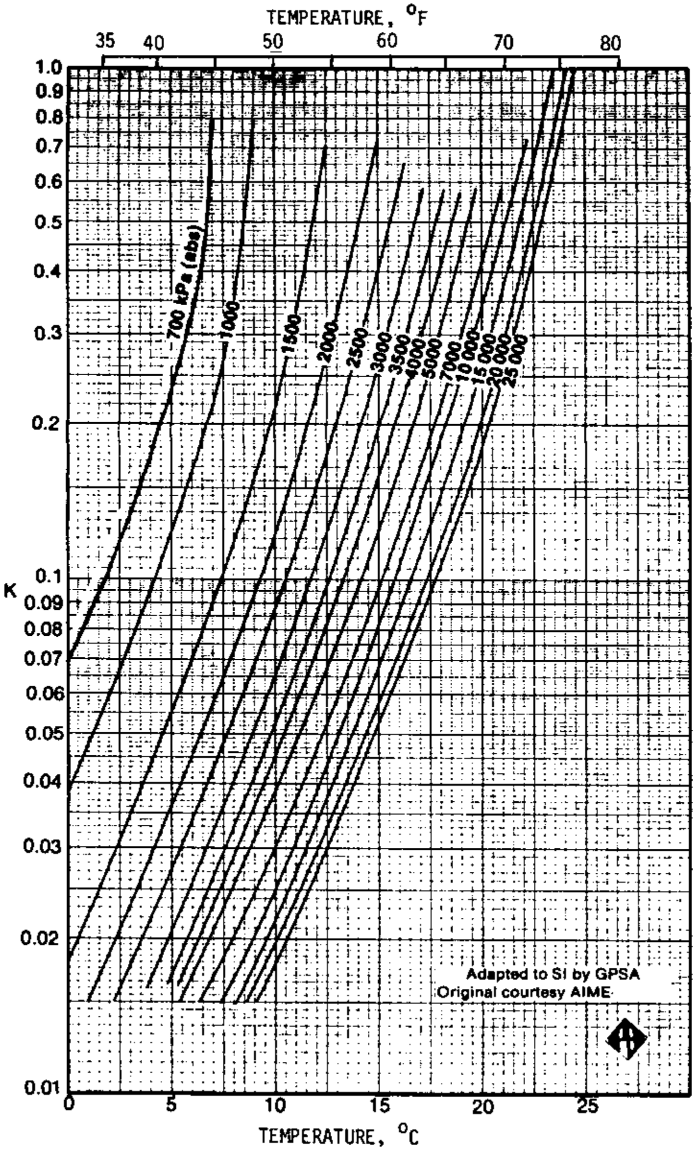


Figure 6.15 Vapor-Solid K Values for Propane

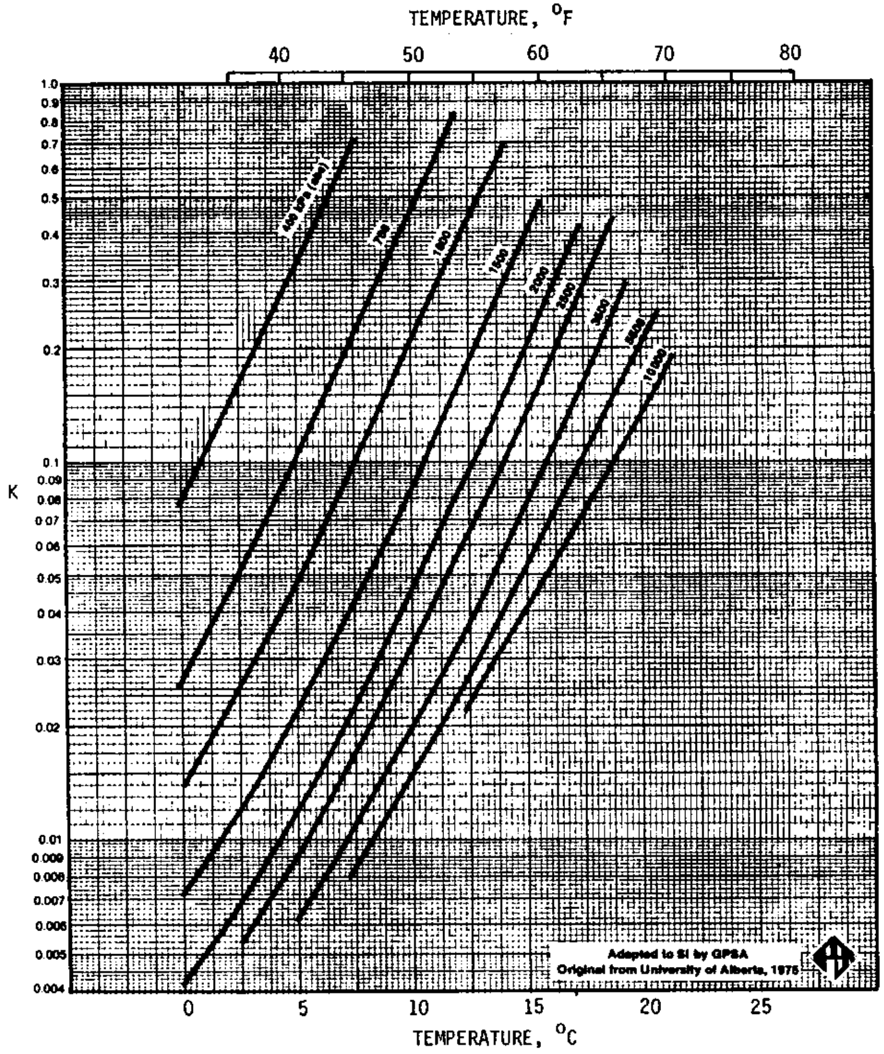


Figure 6.16 Vapor-Solid K Values for iso-Butane

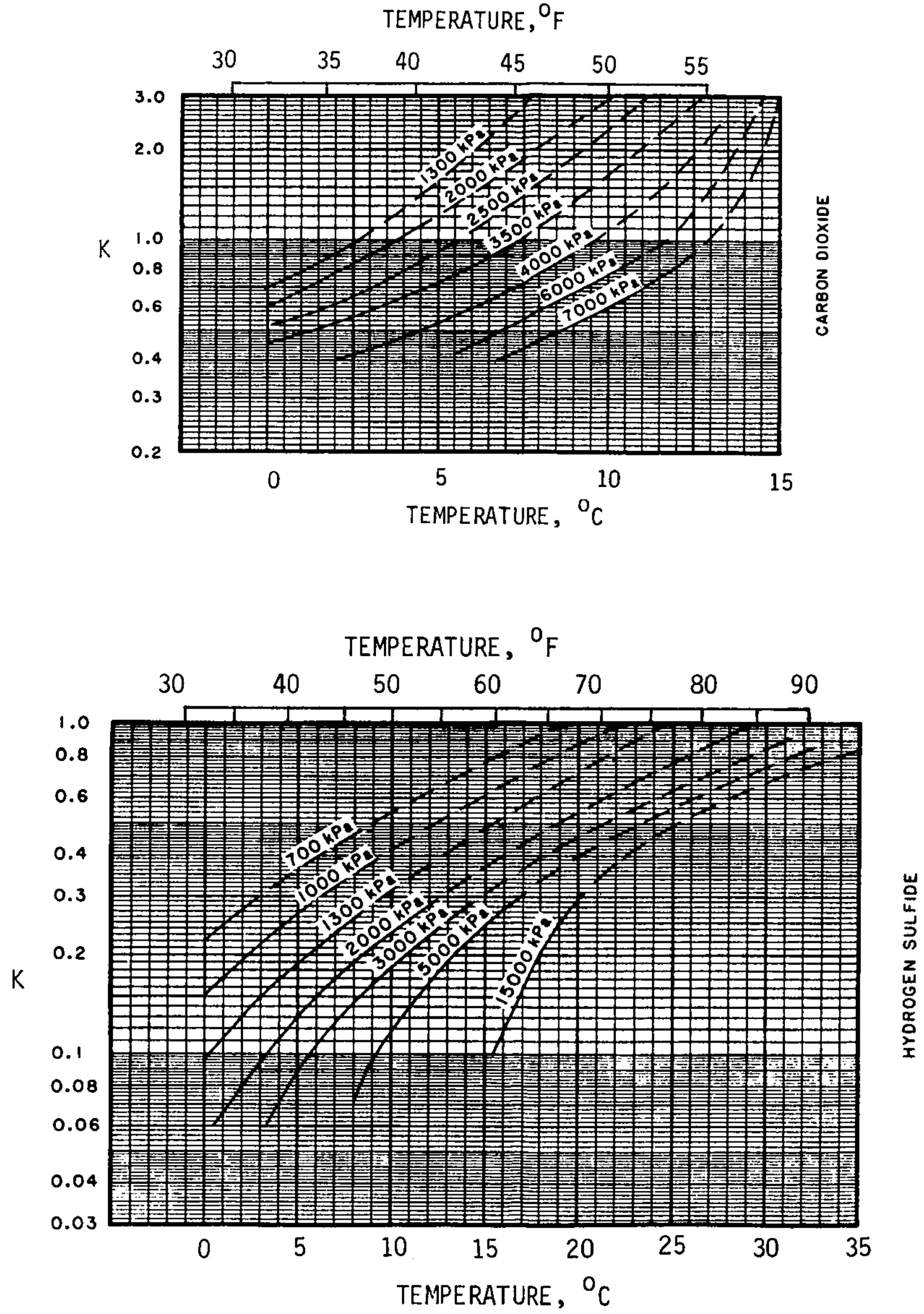


Figure 6.17 Vapor-Solid K Values for CO_2 and H_2S

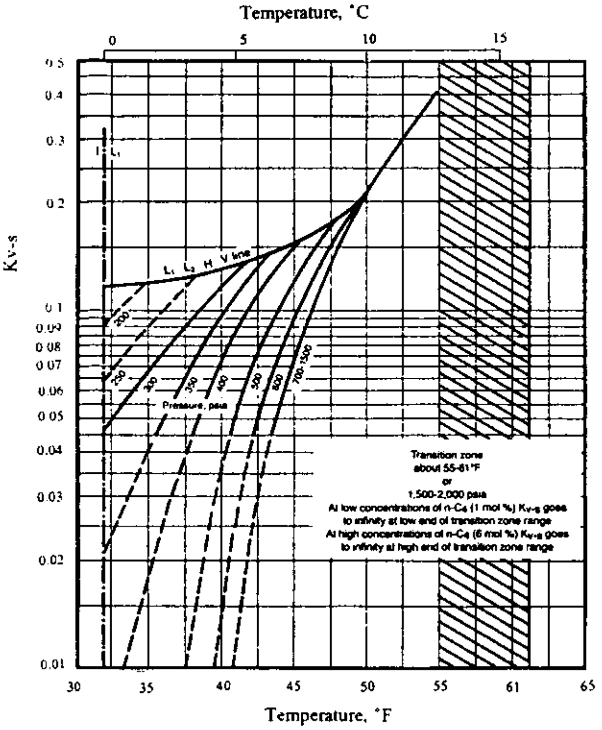


Figure 6.18 Vapor-Solid Equilibrium Constants for n-Butane

A dewpoint type calculation is made using the $K_{v\text{-}s}$ values shown. The term "y_i" is the mol fraction of each component in the gas. $K_{v\text{-}s}$ equals infinity for all nonhydrate formers present.

$$\sum(y_i/K_{v-s}) = 1.0 \tag{6.5}$$

Example 6.3: Find the hydrate formation temperature of the gas below at 2.0 MPa [290 psia].

Kc,12
method

Component	y_i	At 10°C [50°F]	
		$K_{v\text{-}s}$	$y_i/K_{v\text{-}s}$
Nitrogen	0.094	Inf	0.0
Methane	0.784	2.04	0.384
Ethane	0.060	0.82	0.073
Propane	0.036	0.113	0.319
Iso-Butane	0.005	0.047	0.106
n-Butane	0.019	0.21	0.090
CO_2	0.002	2.9	0.001
	1.000		0.974

One can assume a new temperature and repeat the calculation. In Example 6.3 the result will be a hydrate formation temperature of about 9.5°C [49°F].

The preceding approach may also be applied using a different set of $K_{v\text{-}s}$ values. Sloan *et al.*(6.20) present $K_{v\text{-}s}$ values which are dependent upon composition and type of hydrate structure (I or II).

Trekell-Campbell Method

The Katz method possesses pressure limitations and fails to address the hydrate depression effects of molecules too large to fit into the cavities. Too many of these in one location makes it difficult for a stable lattice to form around them.

The molecules larger than methane increase the ease of stable hydrate formation. This method(6.21) uses methane as the reference condition. The additive effect of other molecules is then plotted in Figures 6.19-6.21 for a pressure range 6.9-41.4 MPa [1000-6000 psia]. Each figure is for a specific pressure and the corresponding methane hydrate forming temperature. The abscissa shows the displacement of this methane hydrate forming temperature for the percentage of each component shown on the ordinate.

Notice in Figure 6.19 that n-butane has a slightly positive effect on hydrate formation. In Figures 6.20 and 6.21 the temperature displacement is negative (n-butane becomes a nonhydrate former). This change probably is due to distortion of the lattice with pressure which does not allow even the largest cavities to hold a molecule as large as n-butane.

Figure 6.22 is the negative correction (depression) for nonhydrate formers (pentanes plus). The key parameter is

$$\frac{y_{C_5+}}{1 - y_{C_1} - y_{C_5+}}(100)$$

or the ratio of the mol fraction of pentanes plus to the sum of the mol fractions ethane, propane and butanes. Increasing this ratio is unfavorable to hydrate formation; there are too many large molecules present.

The following general procedure is recommended for this method:

1. Calculate the hydrocarbon dewpoint curve to locate line FEGC in Figure 6.13.
2. Using the 6.9 MPa prediction chart, determine the hydrate formation temperature by algebraically summing temperature displacements. Use the dry gas analysis for the ordinate values and sum the ΔT values found on the abscissa to the temperature for methane hydrate given on the figure. Correct this temperature for the effect of pentanes and heavier if present.
3. Repeat the previous step at 13.8 MPa and correct for pentanes and heavier.
4. Calculate the hydrate temperature at each remaining pressure for which there is a graph until the hydrocarbon dewpoint pressure is reached.
5. Plot the calculated temperatures and fit the best curve through the points.

The procedure shows that pentane and heavier have a calculable effect on hydrate point, and the "K" value is not infinity.

Limited data show that H_2S and CO_2 elevate the hydrate point less in natural gases containing reasonable amounts of ethanes through propanes than would be expected from their binary data with methane. In fact, the correction for CO_2 is negligible for most gases tested. In the final analysis, one does not need an exact hydrate forming prediction. A narrow range of 1-2°C is suitable for design and operation.

I do not recommend that you extrapolate the curves in Figure 6.19-6.22. At some composition the lines approach the vertical (infinite slope) and any increase in concentration produces no further temperature deflection. But, one also must remember that the correlation is empirical and that any extrapolation tends to increase the risk.

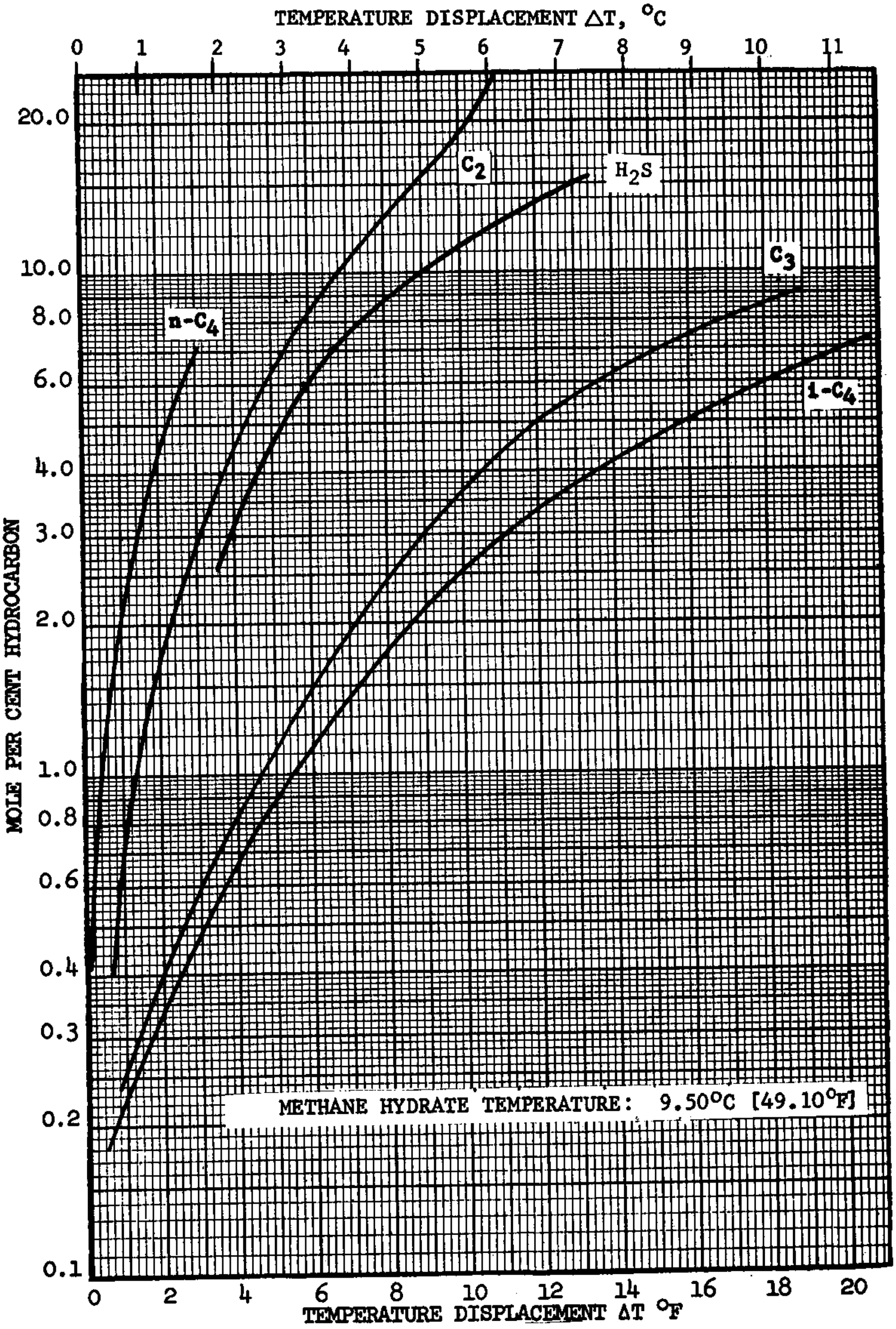


Figure 6.19 Hydrate Prediction Correlations at 6.9 MPa

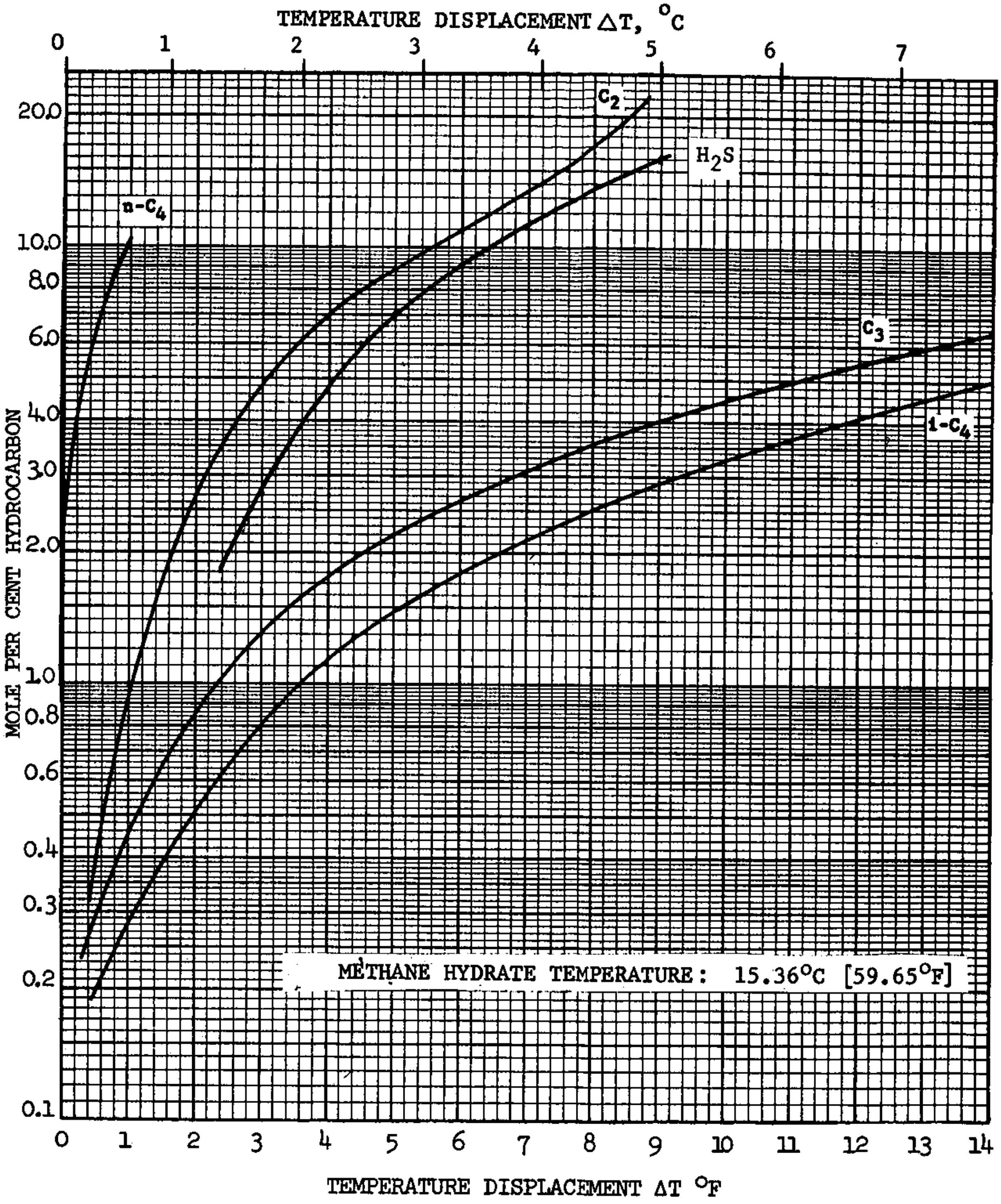


Figure 6.19(a) Hydrate Prediction Correlations at 13.8 MPa

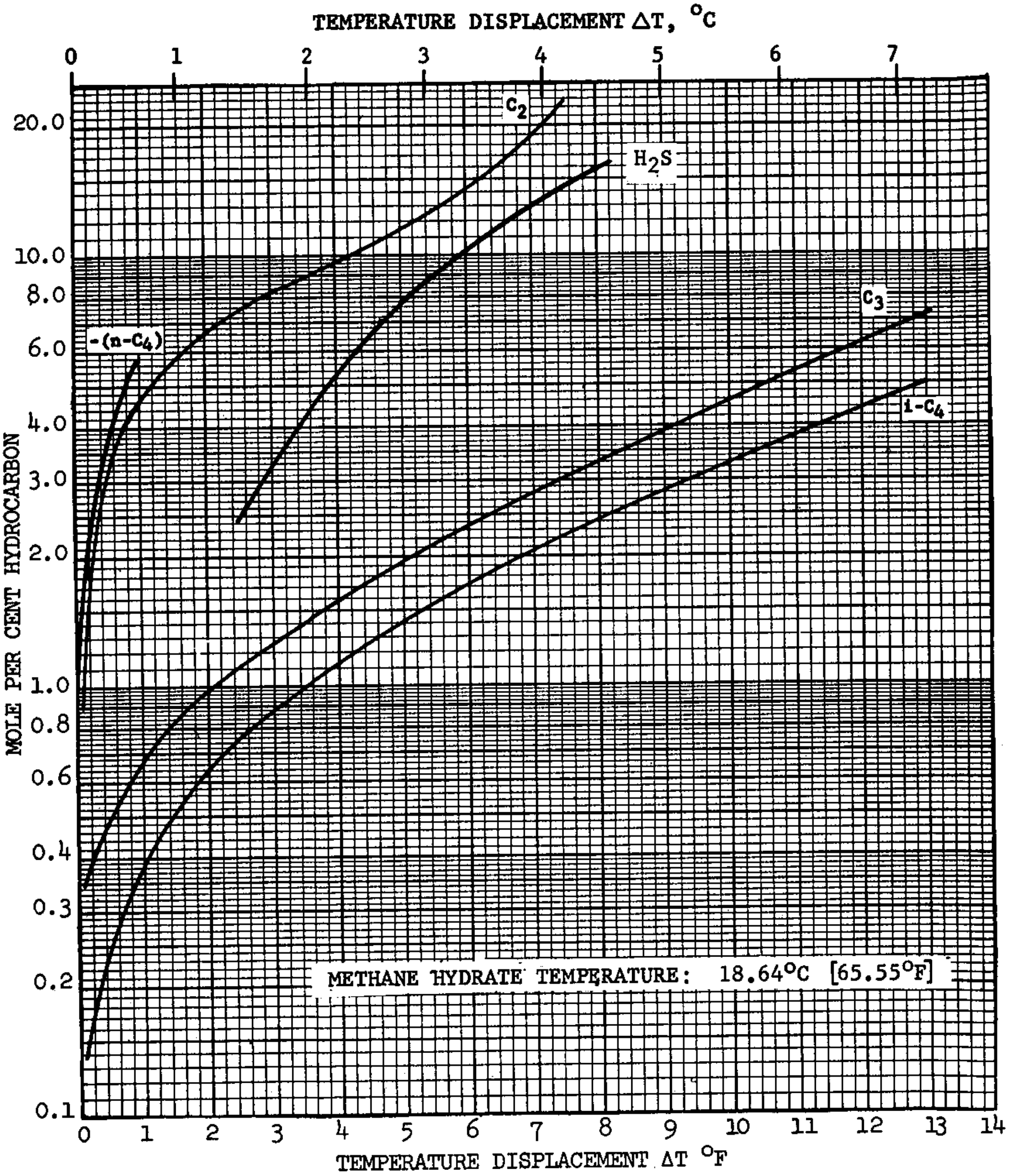


Figure 6.20 Hydrate Prediction Correlations at 20.7 MPa

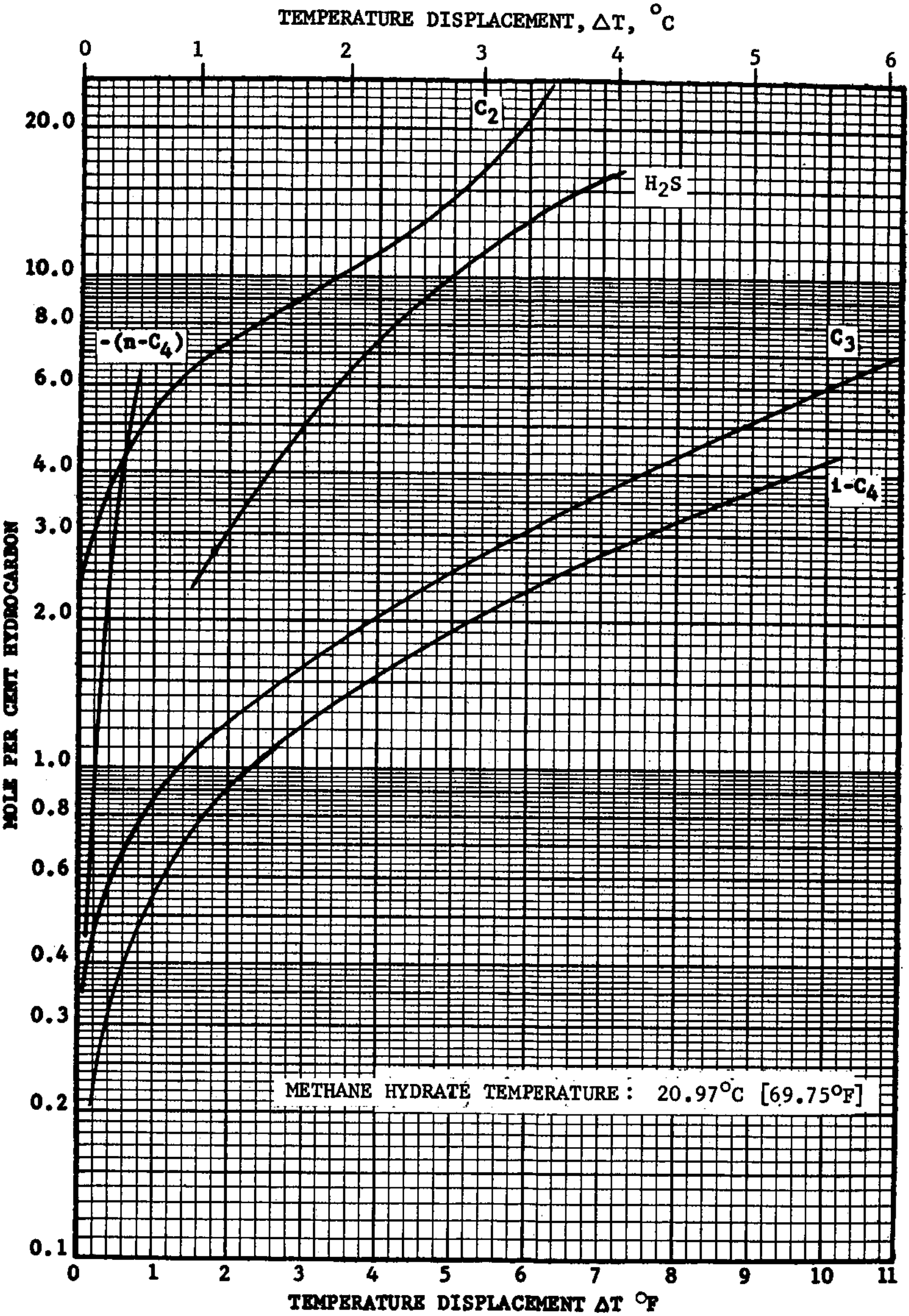


Figure 6.20(a) Hydrate Prediction Correlations at 27.6 MPa

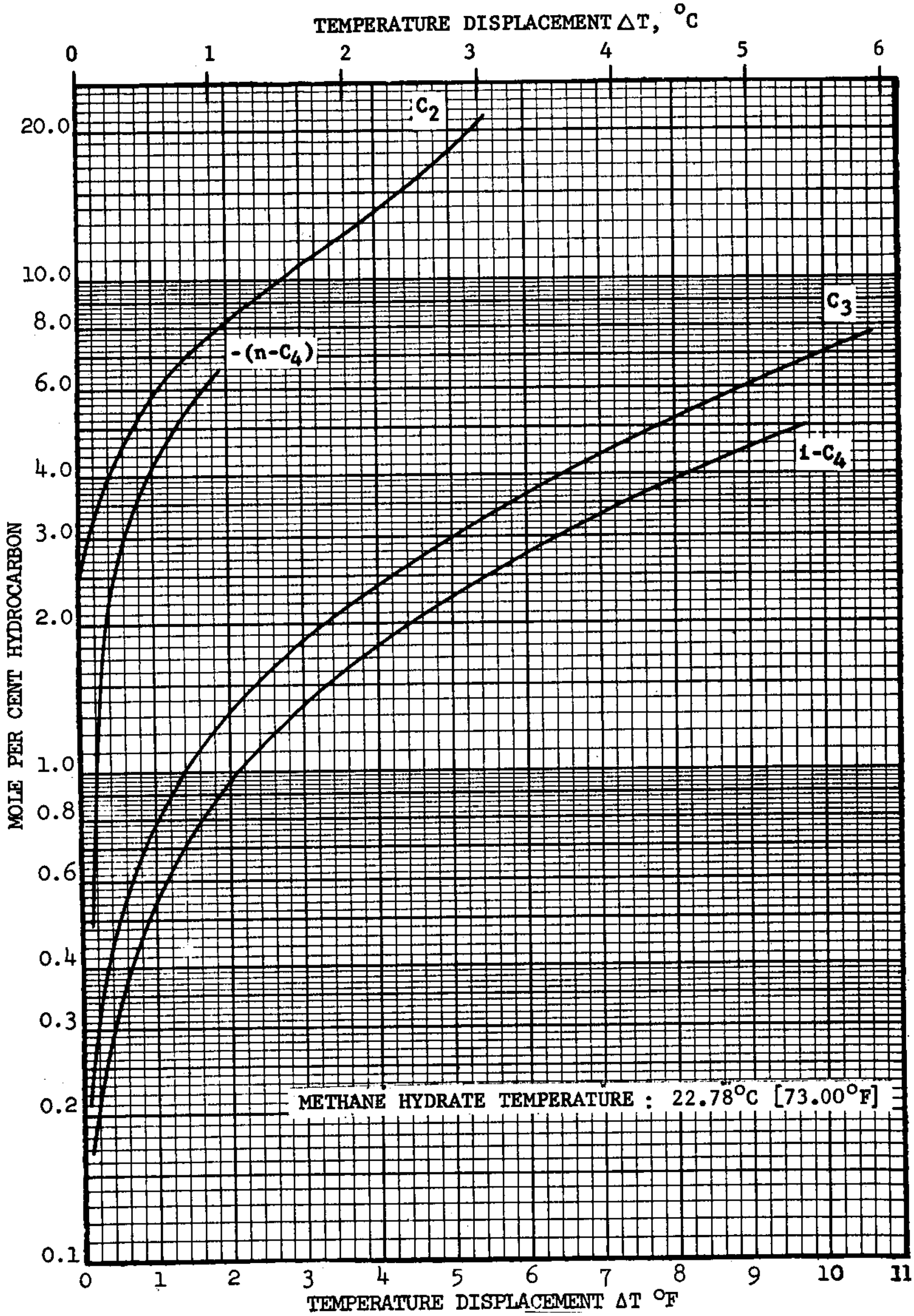


Figure 6.21 Hydrate Prediction Correlations at 34.5 MPa

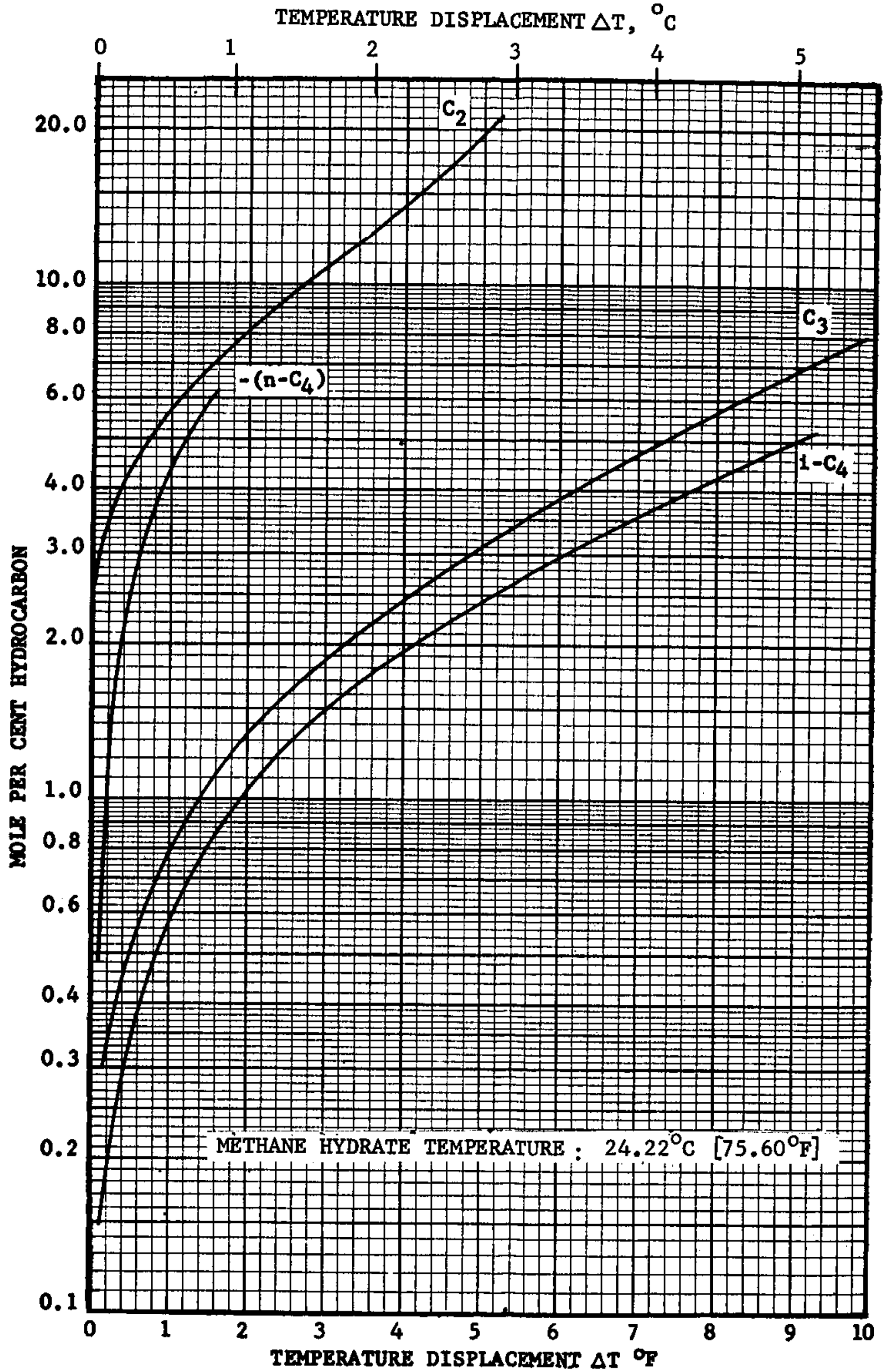
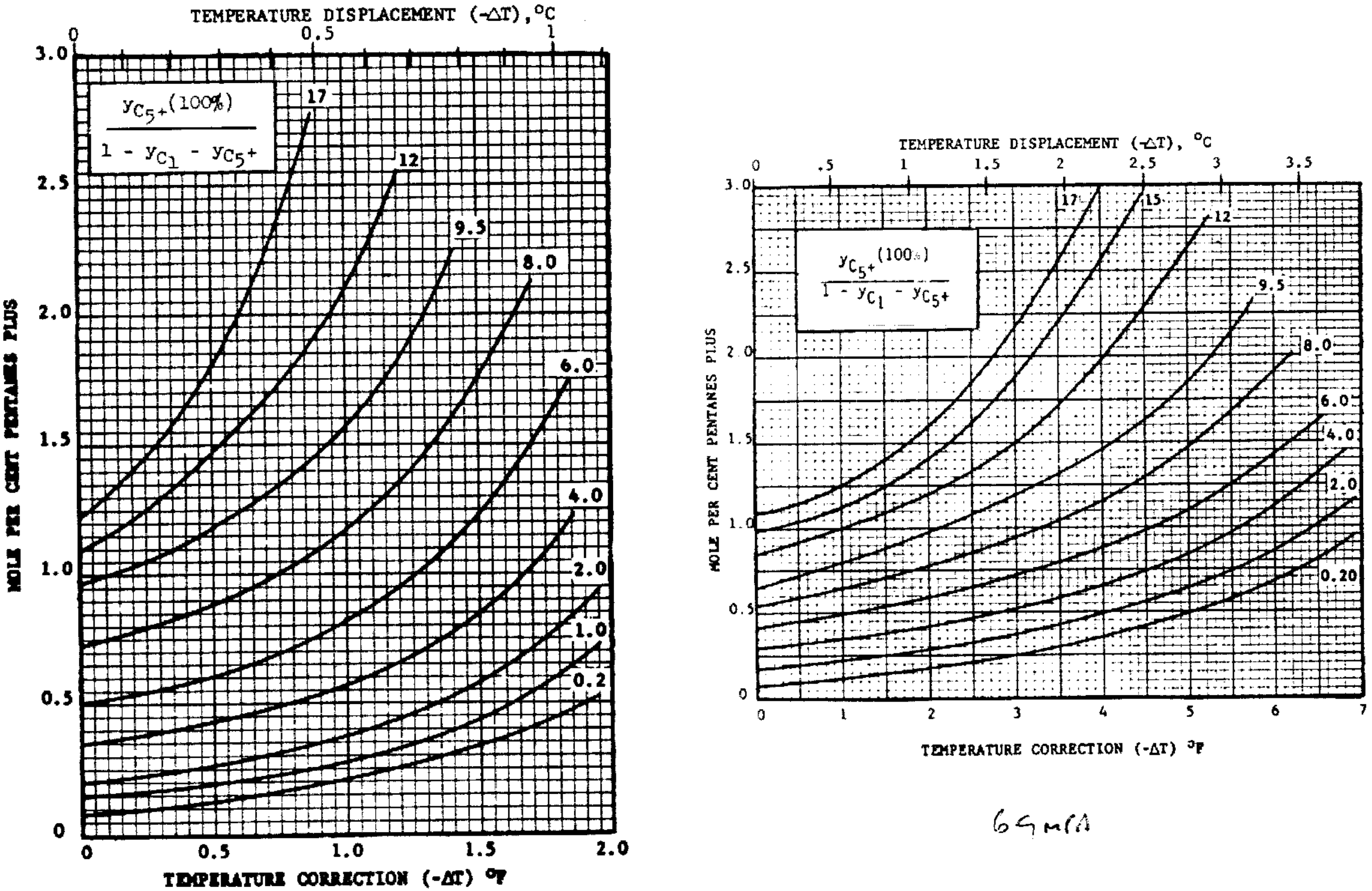


Figure 6.21(a) Hydrate Prediction Correlations at 41.4 MPa



69 MPA

13.8 - 41.4 MPA

Figure 6.22 Hydrate Prediction Correlations for Non-Hydrate Formers

McLeod-Campbell Method

This work(6.22) is for use with high pressure gases above about 35-40 MPa [5000-6000 psia].(6.21) Early work indicated erroneously that gas hydrate forming curves approach the pure methane curve at these high pressures. This work proves that said curves are parallel to the methane curve and have essentially the same slope for all gas mixtures.

A specific correlation was prepared from 41.4-69 MPa [6000-10 000 psia] using a modified Clapeyron equation to describe the energy of phase transformation. Using available data this was reduced to the simple equation

$$T = A\,(C')^{0.5} \qquad (6.6)$$

Where:		Metric	English
	T = hydrate forming temperature	K	°R
	A = correlation factor	2.16	3.89
	C' = component factors from the following table		

Pressure		C Values				
MPa	psia	C_1	C_2	C_3	iC_4	nC_4
41.4	6000	18 933	20 806	28 382	30 696	17 340
48.3	7000	19 096	20 848	28 709	30 913	17 358
55.2	8000	19 246	20 932	28 764	39 935	17 491
62.1	9000	19 367	21 094	29 182	31 109	17 868
69.0	10 000	19 489	21 105	29 200	30 935	17 868

In using Equation 6.6 one expresses the gas composition in terms only of the relative quantity of methane through n-butane. All heavier molecules are ignored. Multiply this pseudo mol fraction by each component's "C" value from the table to find a composite value for substitution into Equation 6.6.

Example 6.5: Calculate the hydrate forming temperature of a gas at 41.4 MPa [6000 psia] with the pseudo analyses shown.

Component	Mol Fr.	C	C′
C_1	0.906	18 933	17 153
C_2	0.066	20 806	1373
C_3	0.018	28 382	511
iC_4	0.005	30 696	153
nC_4	0.005	17 340	87
Total	1.000		19 277

From Eqn. 6.6, $T = (2.16)(19\,277)^{0.5} = 300\text{ K} = 27°\text{C}$ $\qquad T = (3.89)(19\,277)^{0.5} = 540°\text{R} = 80°\text{F}$

Reference 6.23 presents hydrate data for a lean natural gas at pressures to 100MPa [14 500 psia].

Equation of State Calculations

A series of computer solutions have been developed for prediction of hydrate forming conditions.(6.1, 6.24-28) These, unlike the calculations discussed previously are based on equations of state. In general, they are as good as, and maybe better than, the methods discussed above. The models presented here provide a good basis for developing more sophisticated calculations.

APPLICATION OF HYDRATE CALCULATIONS

Remember that hydrates can form only when the gas is saturated with water. This assumption is implicit in any hydrate correlation.

In addition, when using the correlations presented in this text it is assumed that no liquid hydrocarbon or ice is present in the system. (Small amounts of liquid hydrocarbon have little effect on the hydrate temperature.)

Hydrates are a problem only if they are allowed to agglomerate to a degree where they restrict or stop flow. Turbulent flow with liquids present, coupled with sound mechanical design, will reduce the observed hydrate point below that calculated. One should expect all of the above calculation methods to predict hydrate formation temperatures within 1-2 °C of the test value, within the range of their application. In past tests the Trekell-Campbell approach tends to give high values.

One of the best reasons for effective dehydration is to prevent hydrates by eliminating liquid water. Piping, design and valve selection likewise can lower the effective hydrate point. In a multi-phase pipeline the liquid can literally keep the hydrates from attaching themselves and growing. The same is true of wellbore flow. Hydrates may be a problem only when flow stops.

Because of factors like these, hydrate calculations yield somewhat theoretical numbers for planning purposes. If these are used as a proper guideline and the system is designed properly, hydrates should not present a problem.

If you have an unusual gas, outside the normal composition, I recommend laboratory determination of hydrating conditions.

HYDRATE INHIBITION

The positive manner to prevent hydrates (and corrosion) is to keep the lines and equipment "dry" of liquid water. There are occasions (rightly or wrongly) when the decision is made to operate a line containing liquid water. If this decision is made, and minimum line temperature is below the hydrate point, inhibition of this water is necessary.

Many materials may be added to water to depress both hydrate and freezing temperatures. For many practical reasons an alcohol or one of the glycols is injected as an inhibitor, usually methanol, diethylene glycol (DEG) or ethylene glycol (EG). All may be recovered and recirculated, but the economics of methanol recovery may not be favorable in many cases.

Methanol may be used effectively at any temperature. I do not recommend DEG generally below about –10°C because of its viscosity and the difficulty of separation if oil is present. Above –10°C it might be preferred because there is less vaporization loss.

The total injection rate is that needed to provide the necessary inhibitor concentration in the liquid water plus that inhibitor which enters the vapor phase. Any inhibitor in the vapor phase has little effect on hydrate formation conditions.

Equation 6.7$^{(6.29)}$ has been found reliable for prediction of the necessary inhibitor concentration in the water phase to lower the hydrate point a given amount.

$$X = \frac{(d)(M)}{K_i + (d)(M)}(100) \tag{6.7}$$

Where:

		Metric	English
d	= °C depression of hydrate point	°C	°F
X	= weight percent inhibitor in the liquid water phase	–	–
M	= mol wt of inhibitor	–	–
K_i	= constant [illegible]	1297	2335

Equation 6.7 is a simplification of the actual thermodynamic relationship between hydrate depression (d) and inhibitor concentration (X). However, when using the constants shown above, Equation 6.7 does an excellent job of matching laboratory equilibrium data for hydrate suppression with methanol solutions up to about 25 wt% and ethylene glycol to about 50-60 wt%.

Correlating Equation 6.7 with data collected on actual flowing systems is more difficult. On some glycol injection tests the back calculated values for K_i were nearly twice the numbers quoted above.

In truth, no one number applies to all systems, because it is affected by system dynamics, configuration, location and method of injection, etc. Most experienced operators will adjust the injection rate by trial-and-error following initial start-up. This is one of many process calculations that provides little more than a "safe" or "ballpark" estimate to guide operations.

Correlations other than Equation 6.7 have been proposed. For methanol concentrations up to about 50 wt% the following equation$^{(6.30)}$ may be more accurate.

$$d = A \ln (1 - x_m) \tag{6.8}$$

Where:

		Metric	English
d	= °C depression of hydrate point	°C	°F
x_m	= mol fraction of methanol in the liquid water phase	–	–
A	= constant	–72	–129.6

Note that the inhibitor concentration in Equation 6.8 is expressed as a mol fraction rather than a weight percent.

Maddox, *et al.*$^{(6.31)}$ has also proposed a correlation to estimate hydrate depression versus inhibitor concentration. Based on the limited amount of empirical data available, it appears to be more reliable at high inhibitor concentrations (> 50 wt%).

The quantity "d" is found by first predicting the hydrate forming temperature at the maximum pressure in the line segment being protected. The minimum flowing temperature is then estimated. The quantity "d" is the difference between these two numbers. In the absence of a more definitive number, 4°C may be used as a reasonable minimum flowing temperature for buried lines and lines at the bottom of a body of water over 30-40 meters deep.

Total inhibitor injection rate to satisfy the inhibitor concentration needed is found from the equation

$$m_I = m_W\left(\frac{X_R}{X_L - X_R}\right) \tag{6.9}$$

Where:		Metric	English
m_I =	mass of inhibitor solution	kg	lbm
m_W =	mass of liquid water	kg	lbm
X_R =	rich inhibitor concentration	wt %	wt %
X_L =	lean inhibitor concentration	wt %	wt %

The lean inhibitor concentration, X_L, is almost always 100% for methanol and typically varies from 60-80% for glycols. The rich inhibitor concentration, X_R, is determined from Equation 6.7 for field and pipeline inhibition. For inhibition in gas processing plants, the ability to adequately mix the inhibitor and gas in the piping or exchanger is the primary concern. In these cases, the rich inhibitor concentration is set based on two criteria: 1) hydrate depression and 2) mixing. The mixing requirement usually dominates and often results in a specified dilution (X_L to X_R) of 5% or less. The concentrations of both the lean and rich glycol are kept in the non-freezing region (see Figure 6.24).

The following factors are useful for the above calculation.

	MeOH	EG	DEG
Molecular Weight	32	62	106
Density: g/cm^3	0.80	1.11	1.12
kg/m^3	800	1110	1120
lb/ft^3	49.7	69.4	69.6

The injection rate will be in m^3 or liters per unit time in metric units. If water content of gas is found in m^3/10^6 std m^3, this can be converted to liters by remembering that

$$1\ \text{m}^3 = 1000\ \text{liters}\ ,\quad \text{g/cm}^3 = \text{kg/L}$$

One converts the mass rate of inhibitor injection from Equation 6.9 into volumetric rate using an appropriate conversion factor.

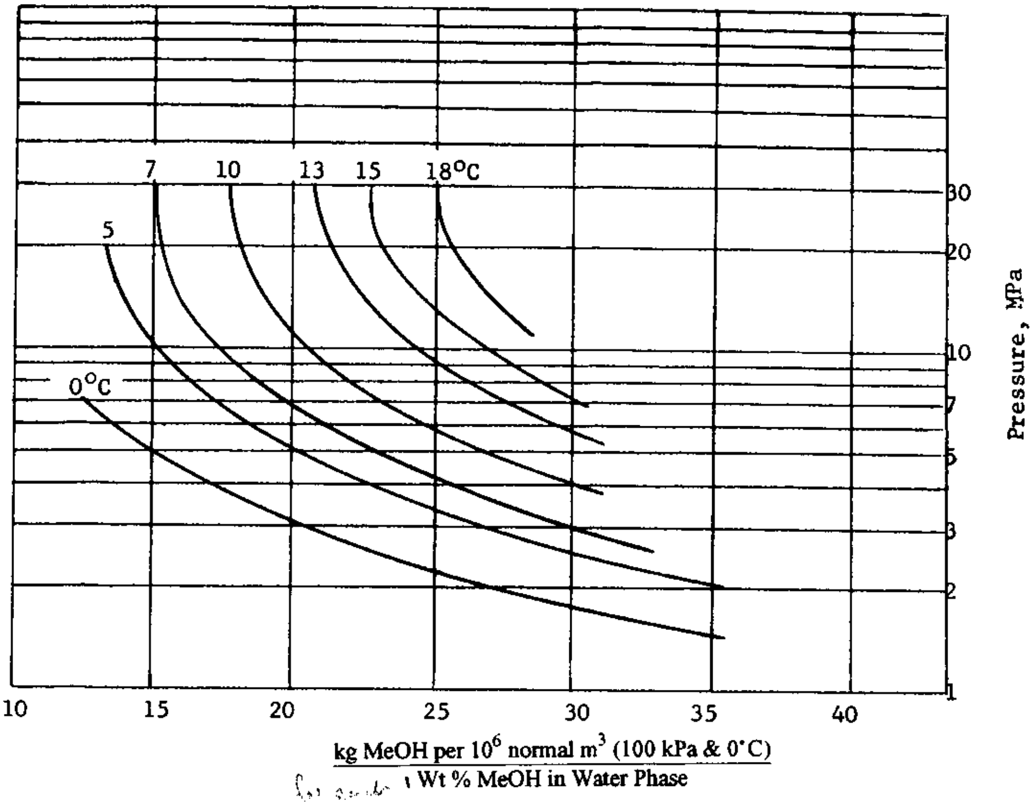
Inhibitor Losses to the Hydrocarbon Phase

Inhibitor may be lost due to its solubility in the hydrocarbon liquid and vapor phases. For glycol systems these losses are small. The following provides useful guidelines.

	Losses	
Vapor	3.5 L/10^6 std m^3	0.23 lbm/MMscf
Liquid	nil	nil

Methanol losses are more significant, particularly vapor phase losses.

Figure 6.23 provides reliable estimates of vaporization loss for pressures less than about 6900 kPa [1000 psia] and water phase methanol concentrations less than about 40 wt%. Enter the ordinate at the minimum pressure of the line segment. Proceed horizontally to the minimum temperature. From this point read vertically to the abscissa, loss per unit volume of gas flowing divided by X, the weight percent methanol in the liquid phase. Knowing X from Equation 6.7 you can solve for methanol vaporization loss per unit volume of gas flowing.



Note: multiply value from chart by 0.95 to convert to std m^3 (100 kPa and 15°C)

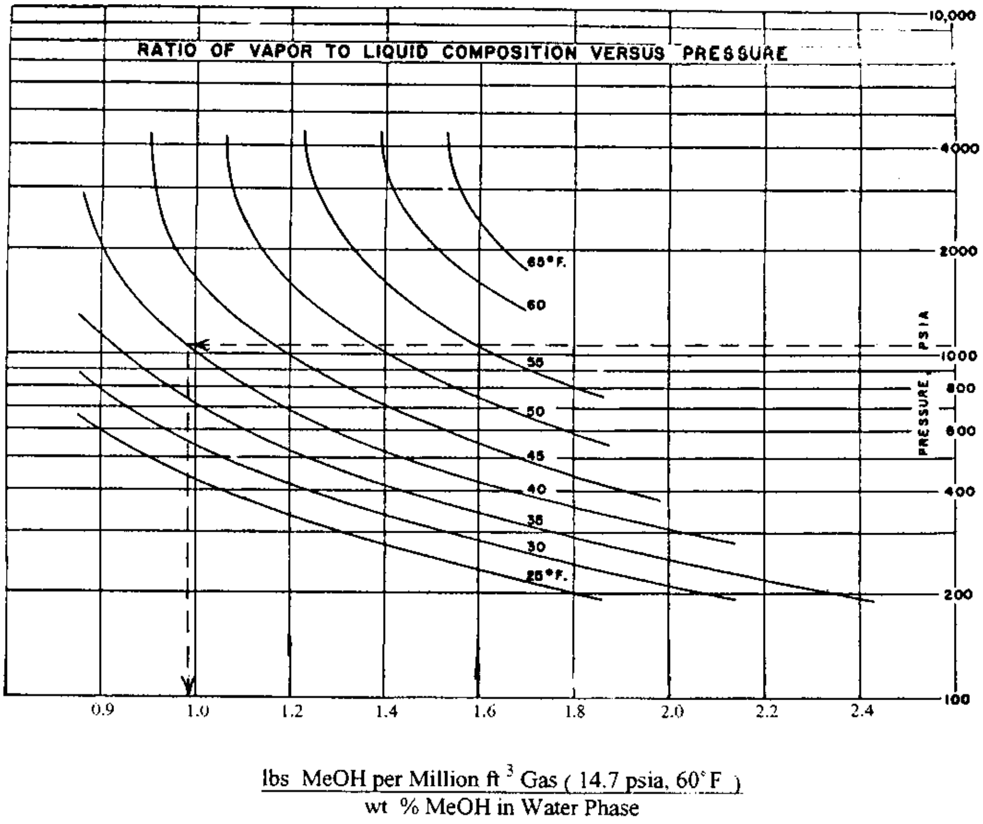


Figure 6.23 Vapor-Liquid Equilibrium of Methanol over Water

At pressures greater than about 6900 kPa [1000 psia] vapor losses may be several times higher than those indicated in Figure 6.23, particularly at high methanol concentrations.(6.33)

Methanol solubility in the hydrocarbon liquid phase is low, however, in systems containing substantial amounts of hydrocarbon liquid the total liquid phase losses can be significant. At typical pipeline inhibition conditions, a solubility of about 0.4 kg/m^3 [0.15 lbm/Bbl] is generally adequate for planning purposes.(6.23) This assumes a paraffinic hydrocarbon liquid. Methanol solubility in aromatic hydrocarbons can be 4-5 times higher than this!

Calculation Summary

The inhibition calculation procedure may be summarized as follows:

1. Determine the hydrate formation temperature of the gas.
2. Establish the lowest temperature expected in the system.
3. Compute the amount of liquid water present at the temperature in Step 2, using the water dewpoint at that temperature and a suitable water content correlation.
4. Use Equation 6.7 to solve for "X." In the equation "d" is the temperature in Step 1 minus that in Step 2. Calculate the weight of inhibitor from Equation 6.9, using X from Equation 6.7. (Note, if Equation 6.8 is used to calculate x_m, it must be converted to a weight percent, X, before being used in Equation 6.9.)

   The volume rate of injection of solution will be the weight of inhibitor per unit time divided by its density, after correcting for concentration.
5. If methanol is used, one must correct for that amount lost to the hydrocarbon vapor and liquid phases. Figure 6.23 is used for this purpose. Enter the righthand ordinate at the pressure at which the temperature in Step (2) occurs. Read horizontally to the lowest temperature and then vertically to the abscissa. The denominator of the abscissa value is the value of "X" previously determined from Equation 6.7 (step 4). Liquid losses may be estimated using a solubility of 0.2 mol %.
6. Total injection rate equals that found from Step 5 plus that from Step 4. This is the rate of inhibitor solution.

Exampble 6.6: 3.5 x 10^6 std m^3/d [124 MMscf] of natural gas leaves an offshore platform at 40°C and 8000 kPa. The hydrate temperature of the gas is 17°C [63°F]. The gas arrives onshore at 5°C and 6500 kPa. Associated condensate production is 60 $m^3/10^6$ std m^3 [10.7 Bbl/MMscf]. Calculate the amount of methanol and 80 wt% EG inhibitor required to prevent hydrate formation in the pipeline.

Methanol:

Step 1 – Hydrate temperature = 17°C

Step 2 – Lowest temperature in system = 5°C
"d" = 17 – 5 = 12°C

Step 3 – Water content @ 40°C and 8000 kPa = 1000 kg/10^6 std m^3

Water content out at 5°C and 6500 kPa = 160 kg/10^6 std m^3

$$\text{Water condensed} = \left(\frac{3.5 \times 10^6 \text{ std m}^3}{\text{d}}\right)\left(1000 - 160\,\frac{\text{kg}}{10^6 \text{ std m}^3}\right)$$

$$= 2940 \text{ kg H}_2\text{O/d}$$

Exampble 6.6 (Cont'd.):

Step 4 – Calculate inhibitor concentration from Equation 6.7

$$X = \frac{(12)(32)(100)}{1297 + (12)(32)} = 22.8 \text{ wt\%}$$

Calculate mass of inhibitor required in water phase from Equation 6.9

$$m_I = \frac{(2940)(22.8)}{(100 - 22.8)} = 868 \text{ kg/d}$$

Step 5 – Calculate losses to the hydrocarbon phases

Vapor – from Figure 6.23 at 4°C and 6500 kPa

$$\text{Losses} \approx 15.7 \frac{\text{kg}/10^6 \text{ std m}^3}{\text{wt\% MeOH}}$$

$$\text{Vapor Losses} = \left(15.7 \frac{\text{kg}/10^6 \text{ std m}^3}{\text{wt\% MeOH}}\right)\left(\frac{3.5 \times 10^6 \text{ std m}^3}{\text{d}}\right)(\text{ 22.8 wt\% MeOH })$$

$$= 1253 \text{ kg/d}$$

Liquid - use 0.4 kg MeOH/m^3 condensate

$$\text{Liquid Losses} = \left(0.4 \frac{\text{kg MeOH}}{\text{m}^3 \text{ condensate}}\right)\left(\frac{60 \text{ m}^3 \text{ condensate}}{10^6 \text{ std m}^3}\right)\left(\frac{3.5 \times 10^6 \text{ std m}^3}{\text{d}}\right)$$

$$= 84 \text{ kg/d}$$

$$\text{Total Injection Rate} = 868 + 1253 + 84$$

$$= 2205 \text{ kg/d} = 0.11 \text{ m}^3/\text{h}$$

For 80% EG solution – calculate X from Equation 6.7

$$X = \frac{(12)(62)(100)}{1297 + (12)(62)} = 36.5\%$$

Calculate mass of inhibitor required in the water from Equation 6.9

$$m_I = \frac{(2940)(36.5)}{80 - 36.5} = 2467 \text{ kg/d} = 0.095 \text{ m}^3/\text{h}$$

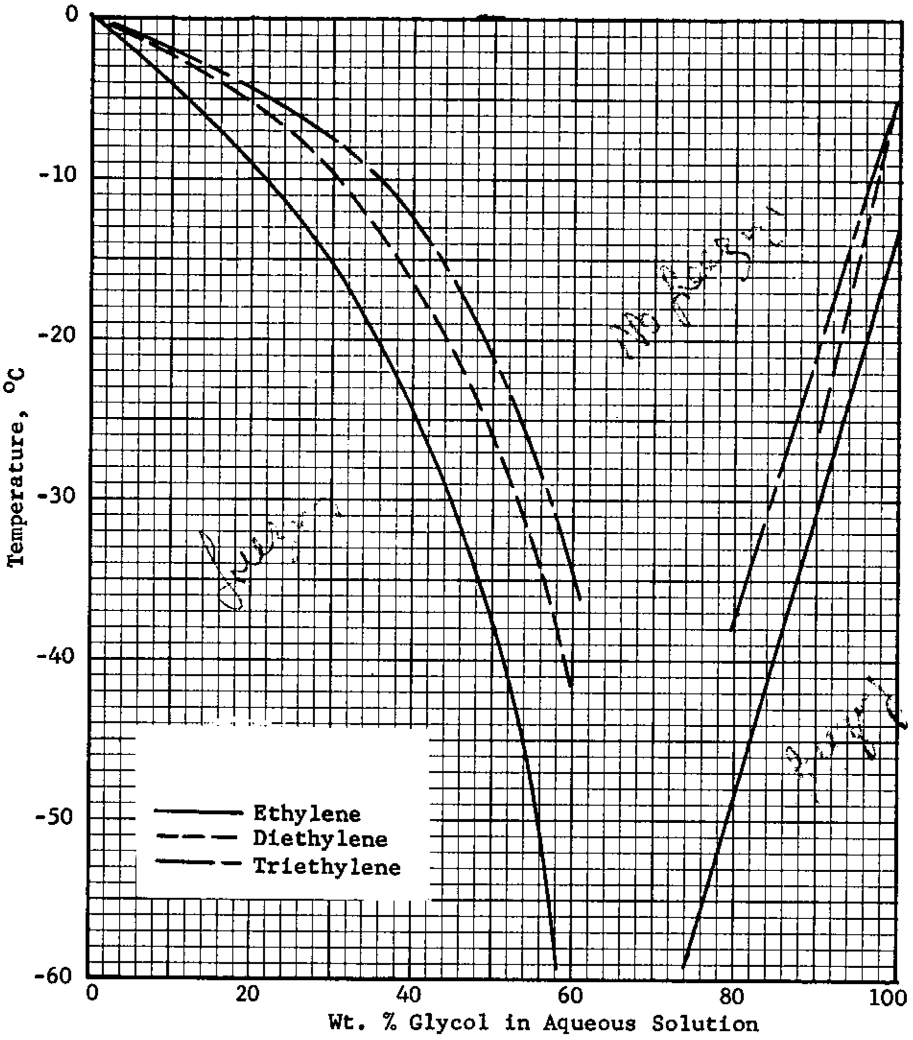
Crystallization (Freezing) of Glycols

Glycols do not freeze solid but form a "mushy" solution that nevertheless does not flow very well. The concentration must be such that this is avoided.

Figure 6.24 shows the "freezing point" for the three most common glycols. The curves go through a minimum. At a given temperature the concentration should be kept between the lines. A value of about 60 wt % glycol is safe at any likely temperature.

If the minimum system temperature is below –10°C [14°F], the injection rate must satisfy both Figure 6.24 and Equations 6.7 and 6.9. Typically the lean glycol solution injected will contain 15-25 wt % water. Thus, the total mount of water to be protected is that in the system plus that injected with the glycol. Equation 6.9 accounts for this water.

With EG and DEG the reboiler should not be operated above 150-160°C [300-320°F] to minimize thermal decomposition.



Benzene
Toluene

MeOH
can also
[illegible]
aromatics
to liquid.

Figure 6.24 Freezing Points of Glycol-Water Solutions

Glycol Losses

The major loss of glycol does not occur from vaporization. It is from losses in the regeneration system, spillage, salt contamination and losses in the separation of oil from the glycol water phase.

Regeneration losses should be small in a well designed unit unless salt contamination tends to plug the still column. Salt is a problem in its own right for the water is distilled off and leaves the salt behind. Salt can be removed from glycol only by vacuum distillation reclaiming.

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7

BASIC THERMODYNAMIC CONCEPTS

Thermodynamics is one of the foundations of all processing calculations, for it furnishes a means by which one may quantitatively characterize a system in terms of measurable variables. This may be done several ways, but the "macroscopic" approach is most convenient and will be used here. This approach examines only the gross characteristics of the system chosen for study, as compared to the "microscopic" one which concerns itself with molecular behavior.

In order to adequately describe the behavioral patterns of our universe, man has had to invent concepts such as pressure, temperature, volume, length, time force, mass, etc. This has been necessary because mere use of our senses (smell, sight, touch, hearing) leads only to percepts such as cold and hot, small and large, push and pull, etc., which involve a large amount of human judgment. What is hot to one person is cold to another. Therefore, the effective communication of ideas, essential to engineering, breaks down because the *percepts* used mean different things to different people.

Thermodynamics is based on rigid definitions of *concepts* so that quantitative conclusions are possible. It is essentially a "language" wherein certain terms have certain meanings. Some of these terms cannot be described in terms of the senses, but in order to be meaningful all must satisfy the following:

1. Be measurable directly or expressible as an exact mathematical function of other measurable concepts or dimensions.
2. Be capable of being described in terms of dimensions and quantities available in our language.
3. Be capable of characterizing a system's state and behavioral patterns so quantitatively that knowledge of the numerical values of the concepts fully describes the system.
4. Mutually vary, one with another, in a systematic way so they may be correlated and thus used conveniently in engineering computations.

Although most of the concepts and dimensions used in our system of thermodynamics are arbitrary (invented by man for his convenience) and not a part of nature, the system has proven so workable and exact that it has led to the development of several absolute laws which all systems follow.

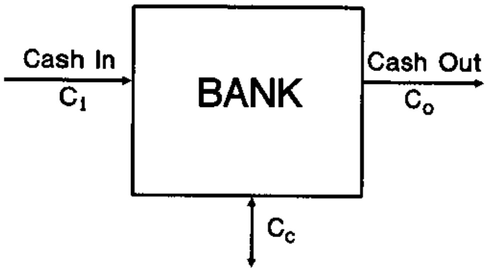
In order to make the system work we must have carefully defined, measurable concepts and dimensions. Two of the most important concepts are temperature and pressure. These are measurable in terms of dimensions established somewhat arbitrarily.

The Accounting Principle

Thermodynamics is the name given to that analytical approach based on the rigorous definition of certain concepts. Standard accounting principles are then applied using these concepts as the input to, or the output from, the system under study. In the event the inputs and outputs are not equal in a given period of time, some provision must be made for a change in the magnitude of the system.

Thermodynamics utilizes concepts that may be related to pressure, volume and temperature (measurable variables) and to each other ... in a systematic manner. The engineer using thermodynamics is an "accountant." He uses mass and energy terms instead of money. A bank account therefore serves as a good illustration of the principle involved.

The summation signs in the following example simply mean that more than one deposit or withdrawal may be made in the time period in question. The term C_c could be plus or minus depending on individual circumstances.



The bank account grows by having a greater cash input than output. The symbol "C_i" represents all withdrawals. The symbol "C_c" represents all inputs and outputs other than deposits and withdrawals – interest, service charges, etc. In a given time period (E – B) the account will change by $C_E - C_B$, where "C" is the size of the account at the beginning and end of the time period of concern. For this system, the following equation applies.

$$C_E - C_B = \sum C_i - \sum C_o \pm C_c \qquad (7.1)$$

Equation 7.1 is absolutely general for all bank accounts, using the terms as defined, if it is recognized that one or more of the terms may be zero during a given time period.

System and Surroundings

In the example, the bank is the *system* chosen for study. The money flows to, and from, the *surroundings.*

This same nomenclature is used in applying the accounting principle to thermodynamic systems. The system is that portion of the universe being studied ... a single piece of equipment or an entire plant. The surroundings is that portion of the universe surrounding the system under study ... the air, ground or water.

Basic Thermodynamic Accounting

Thermodynamics uses the same accounting principle outlined above. It is based on several basic ideas:

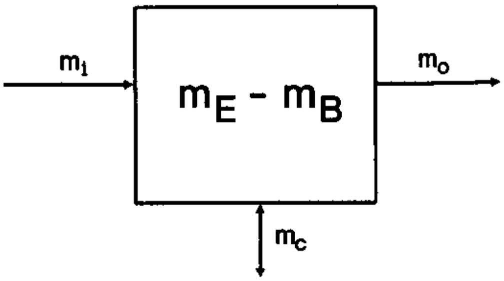
1. A rigid definition of terms capable of accounting for all forms of energy and mass that may possibly enter or leave any system ... across the walls of the equipment or by means of pipe and electrical connections.
2. Development of some basic equations (called laws) that apply rigorously to all systems.
3. The use of concepts that satisfy all four requirements for said concept to have a meaningful application.

Accounting for Mass

The same accounting principle used for money may be used also for mass. The resultant equation is called a mass balance, and may be written as

$$m_E - m_B = \sum m_i - \sum m_o \pm m_c \qquad (7.2)$$

Notice that Equation 7.1 and 7.2 have the same form exactly. The summation sign (Σ) in front of m_i and m_o is used to show that more than one stream may be entering or leaving a given system being studied. These two summation terms simply mean "add up" the mass of all streams entering and leaving.



The term "m_c" is zero unless atomic transformations are taking place within the system; mass may only be created or destroyed by conversion of an equivalent amount of energy.

If there is no accumulation of mass within the system under study, $m_E - m_B = 0$ and it is in a "steady state" condition.

Equation 7.2 is an absolute equation in principle, but not in practice because:

1. We can never measure flow rates without some error.
2. In some systems – like a petroleum reservoir – it is impossible to measure all streams that enter or leave the portion under study (or even be sure what streams enter or leave).

The material balance is somehow reduced to a form where it contains only one unknown. The accuracy of this number depends on how accurately all streams affecting the balance are identified and measured. The final answer must reflect the algebraic sum of all errors.

Energy Balance

The energy balance is not so obvious even though the principle is exactly the same. Energy comes in many forms, many of which we cannot measure directly. But ... it is not necessary to isolate each for consideration so long as all forms are "accounted for."

The intelligent thing is to divide the total energy into a convenient number of subdivisions. What is needed is the least number of divisions whose energy level may be determined quantitatively.

In the classical thermodynamics approach energy may cross the boundary of any system under study in two ways:

1. Associated with the mass entering or leaving.
2. Independent of the mass transfer.

The subdivisions chosen must, of course, be compatible with these two mechanisms so that no energy is divided between them.

This is an energy "sorting" process equivalent to that used in the mails. The rigorous definition provided for each energy subdivision is the "zip code."

The total energy change of the system is the algebraic sum of all energy crossing the boundary of the system during the time period of concern, plus or minus any internal changes due to atomic transformations.

Before writing an energy balance it is necessary to have a means to define total energy entering or leaving the system. This is not routine for many forms of energy can neither be seen nor measured directly. The accepted approach is to define two forms of energy that may be calculated directly – potential (PE) and kinetic (KE) – and then define other forms of energy that may be measured indirectly as mathematical functions of pressure, volume and temperature.

Potential energy is defined generally as the energy possessed by a system as a result of its position in a force field. The primary concern is with gravitational, magnetic and electrical fields. In natural gas processing, gravitational potential energy is the major consideration. For a system whose height above the earth's surface is negligible in relation to the earth's radius, gravitational potential energy (PE) may be found by the following equation

$$PE_2 - PE_1 = \Delta PE = \frac{mg}{g_c}(X_2 - X_1) \tag{7.3}$$

Where: PE = change in gravitational potential energy of the system in time E – B and/or change as a result of the different height of the inlet and outlet streams entering or leaving the system.

$X_2 - X_1$ = change in height of system and/or streams in relation to a datum line (X = 0).

For installations within several thousand feet of sea level "g" and "g_c" are very nearly equal numerically and may be canceled out with little loss of accuracy.

Kinetic energy is that energy possessed by a system by virtue of its motion. If the velocity is negligible when compared to that of light, the mass of the system is essentially independent of velocity an may be considered constant. Velocity has the units of length per unit time and is a means of specifying the relative motion of a body to that of the viewer. It is customary in most process work of the type discussed herein to assume that the viewer has zero velocity, i.e., is standing still. On the basis of these assumptions, KE may be found from Equation 7.4.

$$KE_2 - KE_1 = \Delta KE = \frac{m(v_2^2 - v_1^2)}{2\,g_c} \tag{7.4}$$

Where: v = velocity relative to viewer

m = mass

(All other terms have been previously defined)

The macroscopic velocity of any one stream is found by dividing the volume rate of flow by the cross-sectional area of the channel or pipe. For pumps, compressors, blowers, and the like, changes of kinetic energy are often ignored when calculating work and heat requirements.

Internal energy is a conceptual term defined as all of the energy possessed by a mass other than potential or kinetic. The symbol "U" is used commonly to represent internal energy. By this definition, internal energy is the sum of all forms of energy possessed by a mass other than kinetic and potential.

By this definition of "U,"

$$\text{Total Energy} = U + PE + KE \tag{7.5}$$

Equation 7.5 expresses the total energy possessed by a mass provided some means for calculating internal energy is provided. In a flowing system an additional amount of PV energy is involved. The quantity of fluid entering at the inlet boundary (V_1) can do so only by displacing an equal volume of fluid already there at pressure P_1. Thus, an additional energy, P_1V_1, is involved.

At the exit end the reverse is true and additional energy P_2V_2 is involved. To facilitate the inclusion of these PV terms into the energy balance a new concept, *enthalpy*, is defined by the equation

$$H = U + PV \tag{7.6}$$

Since work (defined following) also may contain a PV term, sometimes there is confusion between work and the PV term of Equation 7.6. In this equation the PV term is an energy change *associated with the mass involved.* Work is non-mass associated.

As noted in later chapters enthalpy is the primary concept used in energy correlations. It is seldom that one has a need to calculate "U" or "PV" separately in Equation 7.6.

The energy terms U, H, PE and KE are sufficient to describe the energy associated with the mass contained in, flowing into or flowing out of, the system. The next step is to define energy terms independent of mass.

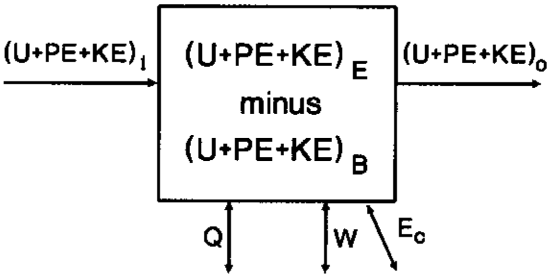
There are two mechanisms by which energy may enter or leave a system without any mass being involved. The are called *work* (W) and *heat* (Q). The total amount of such non-mass associated energy is the algebraic sum of Q and W.

Heat (Q) is defined as all energy crossing the boundary of the system in the prescribed time period that is in no way associated with mass, solely as a result of a temperature difference between the system and the surroundings. Heat gained by the system is arbitrarily defined as plus (+). That lost is minus (–).

Work (W) is defined as all energy crossing the boundary of the system in the prescribed time period that is in no way associated with mass transfer, by any mechanism other than a temperature differential, e.g., all non-mass associated energy crossing the boundary other than that falling in the category of heat. Work done by the system (leaving it) is defined as plus (+). That done on the system (entering it) is minus (–).

Thus a system can gain or lose energy in four ways: (1) energy associated with the mass entering, (2) energy associated with the mass leaving, (3) work and (4) heat.

If we put all this together one may write an overall energy balance using the accounting principle.



$$(U + PE + KE)_E - (U + PE + KE)_B = \sum (H + PE + KE)_i - \sum (H + PE + KE)_o + Q - W \pm E_c \qquad (7.7)$$

Equation 7.7 is called *The First Law of Thermodynamics.*

Specific Applications

Equation 7.7 covers all systems analyzed by man to date. Since some of the definitions are openended, it will most likely cover all systems in the future. However, some terms are zero for a given system. The first practical step is to eliminate from this general equation all terms whose numerical value is zero (or so close to zero as to be negligible in the calculation being made). The following are useful guides in eliminating a term from this First Law equation, to reduce it to the form in which it is applied.

1. $(U + PE + KE)_E - (U + PE + KE)_B$ or $(\Delta U + \Delta PE + \Delta KE)_{E-B}$

   The left-hand side of the equation applies only to the system itself and not to energy associated with mass entering or leaving. If the system is stationary during the time period of concern, ΔKE and ΔPE are equal to zero. The left-hand side then reduces to ΔU.

If there is no accumulation of energy in the system (usually characterized by no accumulation of mass), then U must also be zero. For the case where all three terms are zero the system is called *steady state* – time is not a factor. Since almost all systems discussed in this book basically are steady state systems, the working equations developed for them are based on the left-hand side being zero.

2. $(H + PE + KE)_i - (H + PE + KE)_o$ or $(\Delta H + \Delta PE + \Delta KE)_{i-o}$

   All terms in this part of the equation apply only to the fluid streams entering and leaving. Even though there are elevation changes between inlet and outlet openings, ΔPE is normally small and considered zero for design and analysis. ΔKE is also considered zero for most applications. Velocity changes between inlet and outlet, although measurable, are negligible.

   These assumptions are standard for heat exchangers, pumps, expanders, separation vessels and compressors. About the only process where ΔKE is formally considered would be one in which a deliberate velocity increase is the primary purpose of the device – orifice plate, nozzle, ejector. Unless the potential energy effect on pressure influences the energy requirements, it is ignored.

   If in doubt, calculate ΔPE and ΔKE to be sure.

3. Heat – Q

   Heat is energy in transition across the boundary of the system; it is not a system property. It may be assumed to be zero in the following circumstances:

   a. There is no temperature difference between the system and its surroundings.
   b. The system is properly insulated.
   c. There is insufficient time for heat to be transferred.

   Condition (b) is used for commercially insulated vessels. The heat loss or gain of about 5% or less ignored in the balance is corrected when sizing the equipment. Condition (c) is normally used for steady state flow when the system is a valve or choke and the inlet temperature is near ambient conditions.

   When Q equals zero, the process is called *adiabatic*.

   The plus sign in front of "Q" in the energy balance is based on the arbitrary sign convention that heat entering the system is positive; that heat leaving is negative.

4. Work – W

   Work is also energy in transition across the system boundary. It is zero unless some mechanism is provided for its transfer. The available mechanisms include the following, where energy is transferred without associated mass being transferred:

   a. Pumps, compressors, turbines and expanders connected to some driving, or driven, device (rotating shaft).
   b. A moving boundary in the system (piston in a cylinder, for example).
   c. Electrical wires crossing the system boundary.
   d. Use of a magnetic field to transfer energy.

   In the absence of a device meeting the definition of work, W = 0.

   Work done on the system is negative. Work done by the system is positive. The general energy balance equation is written to reflect this arbitrary definition.

   e. Atomic Transformation – E_c

      None of the systems discussed in this book involve atomic transformation – destroying mass to create energy, or vice versa. Therefore, $E_c = 0$ for all of them.

The Second Law of Thermodynamics – Entropy

The accounting process discussed above is not capable of describing the practical limits of a process, or if it will work at all. It merely keeps track of the energy and mass quantities if the process does proceed.

A third concept has been developed, known as entropy, to characterize how a process must proceed. This concept is like all others; it cannot be perceived directly. It can, however, be expressed as a function of pressure, volume and temperature. It likewise can be described statistically. The many attempts to describe it perceptually have done nothing more than confuse the worth and utility of the term. As a practical matter, it should be regarded as a concept that has proven useful in countless applications.

Entropy is a "yardstick" of the orderliness of a system; the higher the entropy the more disorderly it is. A gas possesses a higher entropy than a solid because its randomly moving molecules are more "disorderly" than those in a solid. Thus, an entropy change as well as an enthalpy change occurs with every change of phase.

This quantity called entropy may be calculated from pressure and temperature. Therefore, a value of "S" may be found for the system and the streams entering and leaving during any given time interval.

Traditionally, in an attempt to give entropy some kind of physical meaning, thermal processes have been used. In such processes only heat is transferred across the system boundary. Therefore, the term (Q/T_b) has been used to calculate the non-mass associated entropy crossing the boundary. T_b is the temperature at the system boundary.

Using the accounting principle and these definitions, the following equation is always representative of an *actual* process.

$$S_E - S_B > \sum S_i - \sum S_o + \sum (Q/T_b) \pm S_c \tag{7.8}$$

A most convenient way to introduce an equal sign into the above equation is to add some quantity to the right-hand side. Let this be some empirical quantity we call *entropy production* and represented by the symbol S_p. The result is reference Equation 7.9. As the above equation would indicate, S_p is always a positive number when applied to an actual system. To date, the application of Equation 7.9 to all actual systems has confirmed this – without exception.

$$S_E - S_B = \sum S_i - \sum S_o + \sum (Q/T_b) + S_p \pm S_c \tag{7.9}$$

Where:
$S_E - S_B$ = entropy change of the system in time period E – B
S_i = entropy associated with all streams entering the system
S_o = entropy associated with all streams leaving the system
T_b = absolute temperature at boundaries of the system
S_p = entropy production in the system
S_c = entropy changes accompanying any atomic transformations
Q = heat transferred to or from the system

Inasmuch as S_p cannot be expressed as a function of measurable variables, except through application of Equation 7.9, it must be regarded strictly as an empirical variable used to characterize the relative efficiency of a system. A numerical value may be found only by determining values of the other terms in Equation 7.9 (or eliminating certain of them) and solving for S_p. In spite of the fact that it does not have the implied physical significance of U, H, and S, which are properties of a system, it is very useful. Calculation of S_p for a process enables one to judge its approach to ideality (its efficiency). Friction loss in fluid flow systems is a function of S_p. If $S_p = 0$ there is no friction loss.

As a practical matter, S_p may be considered a real measure of efficiency. It may be obtained, numerically, from Equation 7.9 by calculating values for all other terms using techniques to be outlined later. $S_p = 0$ represents perfection. The closer the value of S_p to zero, the more efficient the process.

Equation 7.9 is known as the *Second Law of Thermodynamics.* All systems examined to date have conformed to it. Therefore, these First and Second Laws govern the behavior of all systems.

For many common processes like compression and expansion it is inconvenient and impractical to calculate S_p. The standard approach is to assume a *reversible process*...$S_p = 0$. If one further recognizes that a compression or expansion device is steady state, the left-hand side of the Second Law is zero. Since it is impractical to predict "Q" accurately it is assumed to be zero. For all of these conditions, the Second Law reduces to

$$\sum S_i = \sum S_o \tag{7.10}$$

The First and Second Laws of thermodynamics may be combined to yield an equation of the form.

$$\int VdP + \Delta PE + \Delta KE = -(T_b S_p) - W \tag{7.11}$$

Where: T_b = absolute temperature of the flowing fluid

Equation 7.11 is applicable only for the isothermal flow of fluids. It serves as a basis for most of the fluid flow equations used later. For the flow of relatively incompressible liquids such as water, the $\int VdP$ term becomes $V(P_2 - P_1)$.

Relationship Between Thermodynamic Properties and with P-V-T

Macroscopic thermodynamics uses five conceptual property terms, three of which have already been defined. Their defining equations are

$$H = U + PV \quad , \quad G = H - TS \quad , \quad A = U - TS \tag{7.12}$$

Where:
G = derived property called "free energy"
A = derived property called "work function"
etc. = surface energy, chemical energy, and all other forms of energy possessed by a system other than represented by the first two terms of the equation, and potential and kinetic energy

Notice that there is no equation defining "S," even though it appears in several of the above equations. It is practical to think of "S" as that *extensive factor* (dependent on size of system) in the defining equation for internal energy necessary to properly account for temperature effects. The equation for internal energy is written in the form shown because the "etc." is negligible for all processes considered in this manual. In adsorption processes, where surface energy plays a role, we can get around having to include the use of the "etc." term by writing an energy balance which includes the "heat of absorption" and/or the "heat of desorption." Elimination of the "etc." term simplifies the mathematics.

Free energy (G) and work function (A) are important derived properties of a system but have little direct use in the processes covered in this manual.

In actual practice it is easier to write the basic equations in differential or finite difference forms. As it turns out, we are always more interested in the change of U, H, S, G and A rather than their absolute values. For example,

$$dH = dU + PdV + VdP$$

$$\Delta H = \Delta U + \Delta(PV) = \Delta U + (P_2V_2 - P_1V_1) \tag{7.13}$$

$$dU = TdS - PdV + \text{etc.}$$

The above equations make it possible to relate H, U, and S to each other and to P, V, and T. For example, for the above

$$dH = TdS + VdP \tag{7.14}$$

when the "etc." term is negligible. This is a very useful equation.

It is convenient to express enthalpy and entropy data in terms of some "base condition" at which H and S equal zero. Many tables of data for water and steam, for example, arbitrarily assume H = S = 0 for saturated liquid water at 0°C. The values read for any condition are therefore not the absolute value in relation to absolute zero but the relative value above or below the base condition. This approach is convenient and accurate. When using data from two sources for the same substance, it is necessary that they use the same base condition or their values must be adjusted to the base conditions before subtracting them.

Many tables of data list only values of h, s, and v as functions of temperature (T) and pressure (P). Corresponding values of U must be found from Equation 7.13.

H, S, and U are derived properties of a system and are not measurable directly. They may be expressed conveniently in terms of the measurable properties P, V, and T. The following equations have been developed to accomplish this.

$$dH = C_p dT + \left[V - T\left(\frac{\delta V}{\delta T}\right)_p \right] dP \tag{7.15}$$

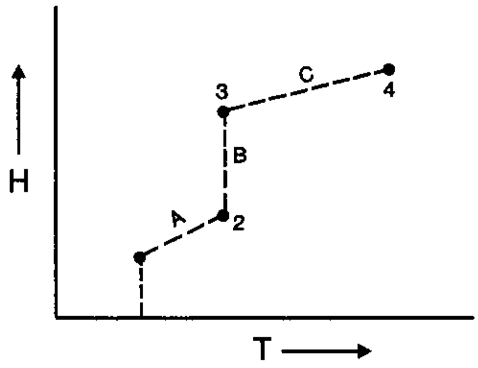
$$dU = C_v dT + \left[\left(\frac{\delta P}{\delta T}\right)_v T - P \right] dV \tag{7.16}$$

$$dS = \frac{C_p dT}{T} - \left[\left(\frac{\delta V}{\delta T}\right)_p \right] dP \tag{7.17}$$

Where: C_p = heat capacity at constant pressure, i.e., $(\delta H/\delta T)_p$
C_v = heat capacity at constant volume, i.e., $(\delta U/\delta T)_v$
P, T = pressure and temperature expressed in absolute units
V = corresponding volume of system expressed as total volume or volume per unit mass

Equations 7.15-7.17 are valid for any process, or portion thereof, for which H, U, and S are continuous functions of P, V, and T, and where no chemical reaction is involved. Mathematical continuity is satisfied so long as there is no change of phase of the system under study. A phase change represents a discontinuous process. Equations 7.15-7.17 can be used only for those portions of a process where no phase change occurs.

H, U, and S are also *point properties* of a system, e.g., exact functions of P, V, and T. This means that the ΔH, ΔS, and ΔU between any two points in a process is dependent only on the end conditions and not on the path followed between these end conditions. It is this characteristic which makes H, U, and S so valuable. One can devise any stepwise process between the end conditions to suit his calculation convenience. The algebraic sum of the ΔH, ΔS, and ΔU changes for each step is equal to the total change between the specified end conditions.



Consider the process occurring in a steam boiler where entering cold water is converted to superheated steam. It may be shown schematically as follows, if it is assumed that pressure is constant in the boiler.

Step A. Heating cold water to saturation temperature

Step B. Vaporizing water to steam

Step C. Superheating steam

From the above, $\Delta H_A + \Delta H_B + \Delta H_C = \Delta H$ (total).

The last term of Equation 7.15 disappears, for at constant pressure dP = 0. None of Equation 7.15 or 7.17 may be used for Step B, since a phase change is occurring. The ΔH for this step will be found by methods to be shown in Chapter 8. The C_p values needed are readily available in the literature.

Special Case for Ideal Gases

For the special case where the assumption can be made that the ideal gas concept satisfies the gas PVT relationship, i.e., PV = nRT, Equation 7.15 and 7.17 reduce to the following:

$$dH = C_p dT \quad , \quad dU = C_v dT \quad \text{and} \quad dS = \left(\frac{C_p dT}{T}\right) - \left(\frac{R}{P}\right) dP \tag{7.18}$$

For ideal gases, $C_p - C_v = R$. The following may be used to estimate C_p and C_v in the absence of specific values for the given gas:

Type of Ideal Gas	C_p	C_v
Monotomic	5/2 R	3/2 R
Diatomic	7/2 R	5/2 R
Triatomic	9/2 R	7/2 R

The value of the gas constant "R" used in the table is normally 8.314 kJ/(kmol·K) [1.987 Btu/lb-mol-°R]. Therefore, C_p and C_v would have the same units. When available, published specific heat values for the gas in question should be used in preference to the above values.

Special Case for Incompressible Liquids

In most of the design calculations of the type presented herein, it is satisfactory to assume that the liquid is incompressible, i.e., v or ρ are substantially independent of pressure and temperature. For this case, Equation 7.15 and 7.16 reduce to

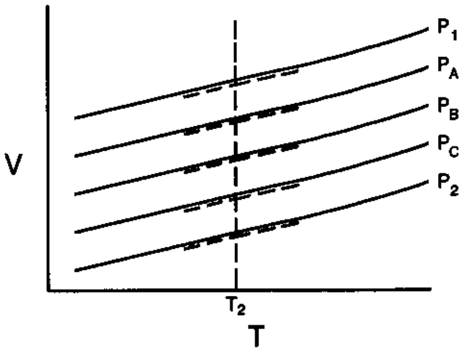
$$dH = C_p dT \quad , \quad dU = C_v dT \tag{7.19}$$

Solution of Equations 7.15 and 7.17

Equations 7.18 and 7.19 are easy to solve once one has values of C_p and C_V. Equations 7.15 and 7.17 are more difficult for non-ideal gases. They are most easily solved by numerical techniques when one has a computer available, after choosing an equation of state to express the PVT relationship of the gas. The principles involved are illustrated below.

When thermodynamic data are available in graphical or tabular form expressing H, U, and S as a function of P, V, and T, a simple graphical solution will suffice. A differential is simply the slope of a line while an integral is an area under a curve.

For example, suppose one wished to calculate the ΔH for compression of a gas from P_1, T_1 to P_2, T_2. He could prepare the following figure.



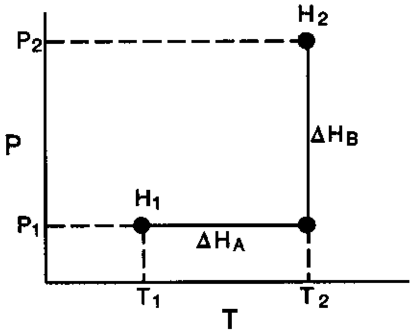
P_A, P_B, and P_C are any pressures between P_1 and P_2. Each slope at T_2 is $(\delta V/\delta T)_p$.

The compression process may be divided into two steps – one at constant temperature and one at constant pressure.

$$H_2 - H_1 = \Delta H_A + \Delta H_B = \Delta H$$

$$\Delta H_A = \int_{T_1}^{T_2} C_p dT$$

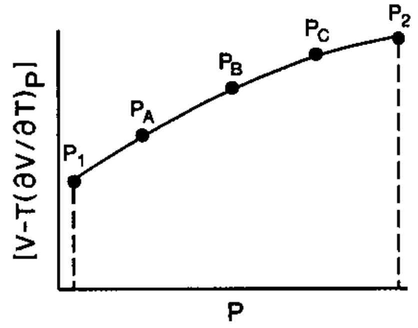
$$\Delta H_B = \int_{P_1}^{P_2} \left[V - T_2 \left(\frac{\delta V}{\delta T} \right)_p \right] dP$$



ΔH_A is found from C_p data. ΔH_B is found from taking slopes of the curves as shown on the V vs. T plot. At each pressure subtract the slope $T(\delta V/\delta T)_p$ from V. A plot of this difference versus P yields the following type plot.

Numerical value of area under curve equals ΔH_B.

The area under the curve may be found using the Simpson or Trapezoidal Rules, with a planimeter, or numerically.



Effect of Pressure on C_p

Most available C_p data are reported at atmospheric pressure. These may be converted to higher pressure using the equation

$$C_p = C_p^o - T \int_{P_o}^{P} \left(\frac{\delta^2 V}{\delta T^2} \right)_p dP$$

Where: C_p^o = heat capacity at atmospheric pressure

In later sections correlations will be presented, based on this equation, which are useful for predicting heat capacity at various pressures and temperatures.

General Effect of PVT Relationships

The foregoing shows that the derived energy terms like enthalpy, entropy and internal energy may be expressed quantitatively as functions of P, V, and T. Later discussions will show that density, viscosity, water content, equilibrium relationships and the like also may be expressed as functions of P, V, and T.

Since all of these quantities are functions of the same measurable variables (P-V-T), they may be intercorrelated. However nebulous the origin of the derived energy terms might be, their relatability to the measurable terms serves to validate their practical value.

Most current computer programs use PVT equations to calculate both the derived and the physical properties of systems.

APPLICATIONS OF THERMODYNAMICS

The foregoing sections summarize the definition of terms and principles in thermodynamics. Some of these are somewhat arbitrary because this is nothing more than an energy inventory control program; to identify and keep track of energy, in all its forms, using concepts that meet the four conditions set forth at the beginning of this chapter.

Chapter 8 outlines the basic methods used to find numerical values for quantities like H and S. These range from simple, direct correlations for pure components to the use of PVT equations of state for mixtures. Volume 3 of this series expands on this discussion.

It will be noted that a change in system energy must occur with every phase change. These are inseparable variables. If a phase change is caused by a change in pressure, temperature or composition as shown in Chapter 4, a corresponding energy change must occur. A controlled energy input will cause a change of P, T or phase. Remember: P and T are indirect measurements of molecular energy. If at saturation these do not change with energy level changes, the phase must change. Many practical design and operating problems are a result of a failure to recognize the inflexible relationship between energy level and phase.

In most of the systems discussed herein, Equations 7.7, 7.9 and 7.11 are simplified before application. Most often Equation 7.7 reduces to $\Delta H = Q - W$. Most applications of Equation 7.9 are for an *adiabatic reversible* (isentropic) process wherein this equation reduces to $\Delta S = 0$.

Expansion across a valve or choke. Since Q and W are zero (or essentially so), Equation 7.7 reduces to $\Delta H = 0$. There is no change in H in this *isenthalpic* process. As shown in Chapter 17 on refrigeration processes, this is the basis for temperature drop and liquid recovery predictions when expanding across a valve.

Pumps, compressors and expanders. The power required (or available) is calculated from Equation 7.7 or 7.11. Most computer programs use the former. At inlet conditions (1) values of H, and S, are calculated. Outlet S_2 is set equal to S by assumption of an isentropic process; a corresponding H_2 can then be found. This theoretical or reversible ΔH is then corrected with an efficiency factor to find an actual ΔH which is equal to work (because $Q = 0$ for an isentropic process).

As shown in chapters 15 and 16, an alternative approach uses a version of Equation 7.11 and what is called a polytropic process to indirectly introduce an efficiency factor.

Heat exchange. Equation 7.7 reduces to $\Delta H = Q$. One calculates ΔH for the process which establishes the heat load from an equation like 7.15, from latent heat correlations (for phase changes) or directly from an enthalpy correlation or calculation model.

Notice in all of the above examples that the basic calculation is always an enthalpy change. One of the thermodynamic equations then is used to relate this change to the quantity desired. This is the most convenient way to make these calculations because enthalpy depends only on the inlet and outlet conditions and is independent of the means by which the flowing stream changed P, T or phase between these two points.

NOTES:

8

CALCULATION OF SYSTEM ENERGY CHANGES

The principles and equations of Chapter 7 plus phase behavior provide the basis for the calculation of the energy required or available at a specified set of conditions. The majority of all operations involves the addition or removal of heat or work. The first law of thermodynamics expresses the concept

$$\text{Energy Accumulation} = \text{Energy Inflow} - \text{Energy Outflow} \qquad (8.1)$$

Equation 7.7 from the previous chapter reduces to

$$\Delta H = Q - W = H_2 - H_1 \qquad (8.2)$$

for the usual assumption that the process is steady state (no energy accumulation in the system) and that potential and kinetic energy changes are negligible.

It is convenient to calculate energy terms *per unit of mass*, written as lower case letters s, h, u, etc. Equation 8.2 would then be written

$$m\,(\Delta h) = Q - W = m\,(h_2 - h_1) \qquad (8.3)$$

Where m = the flow rate in mass per unit time.

In the application of Equations 8.1-8.3, the convenient assumption is to eliminate either Q or W; then correct the results, if necessary, using an "efficiency."

In an ordinary heat exchanger, no work is done so it is satisfactory to assume that $\Delta H = Q$.

In expansion across a value, the amount of heat lost is trivial and there is no work done. So...$\Delta H = O$.

With pumps, the value of Q is normally very small so it is normally assumed that $\Delta H = -W$.

With compressors and expansion turbines a more arbitrary assumption is required to simplify the calculation using enthalpies to obtain $\Delta H = -W$. There are several approaches but the most common one is to assume the process is adiabatic and reversible. The adiabatic assumption removes Q from the equation and the reversible assumption enables one to calculate the theoretical outlet enthalpy. An appropriate efficiency based on operating tests is then used to correct the calculation back to some semblance of reality.

CHOICE OF SYSTEM

The first decision is choice of the system. It may be a single piece of equipment, several pieces, or an entire module. The choice must be based on two primary considerations: (1) what you wish to determine and (2) what data are available. As later correlations will show, if the fluid composition and flow rate are known, you need only to define two thermodynamic properties to calculate the corresponding enthalpy change.

Figure 8.1 is a system to pump and heat crude oil to a higher pressure and temperature. At the pump inlet the P, T, oil composition (or specific gravity), and flow rate must be specified. At the heater outlet P and T must be specified. It would be assumed that the flow rate is the same at both points (1) and (2), i.e., there are no leaks. One can calculate H_1 (h_1) and H_2 (h_2). But this leaves two unknowns in Equations 8.2 or 8.3, Q and W. The best way is to divide this system into two parts – the pump and the heater. The pump will be calculated first (Chapter 14) assuming an adiabatic process (Q = 0). So, W is calculated from the enthalpy change across the pump. Since W = 0 in the gas fired heater, Q equals the enthalpy change from the pump outlet to the heater outlet.

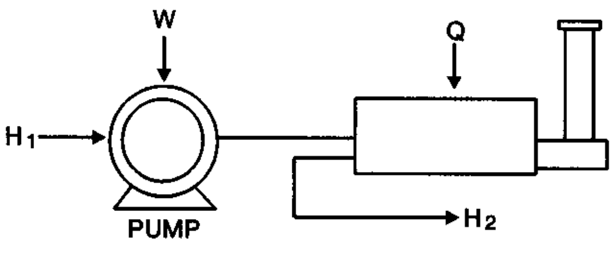


Figure 8.1

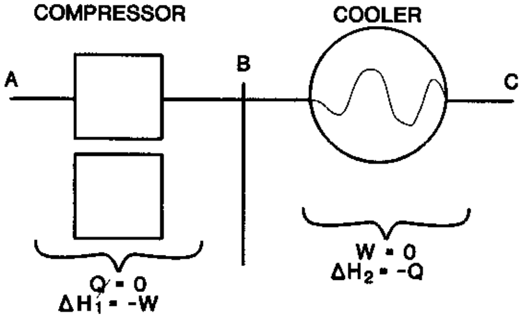


Figure 8.2

Figure 8.2 for a compressor and aftercooler involves the same considerations. In preparing for a calculation, make a sketch of the equipment involved. Write down known (or specified) conditions. Is enough known to make the desired calculation or are further specifications (or assumptions) necessary? Do not make arbitrary assumptions. Use judgmental assumptions compatible with experience with such equipment and local considerations.

UNITS OF MEASUREMENTS

The basic energy units used are kilojoules (kJ), megajoules (MJ) and British Thermal Units (Btu). Power, the time rate of transferring energy, usually is expressed as kilowatts (kW), megawatts (MW) or horsepower (hp).

Specific enthalpy (h) is used most often as kJ/kg or Btu/lbm; specific entropy is kJ/(kg·°C) or Btu/(lbm-°F).

1 kJ = 1000 J = 0.001 MJ = 0.948 Btu, 1 Btu = 1.055 kJ

1 hp = 2685 kJ/h = 2545 Btu/h = 0.746 kW = 550 ft-lbf/s

1 Btu/lbm = 2.326 kJ/kg, 1 Btu/(lbm-°F) = 4.187 kJ/(kg·°C)

1 kW = 1.34 hp = 3600 kJ/h = 341 Btu/h

Flow Rates

Flow rates expressed in volume units must be converted to mass units to calculate total energy transferred per unit time. Appendix 1C (Chapter 1) and Appendix B at the back of the book summarize these factors involved.

Herein we are using the following standard conditions to express gas volumes: Metric 15°C and 100 kPa; English 60°F and 14.7 psia. The conversion factors shown following apply for these standard conditions.

Gas	Liquid
kmol/h = 1739 (10^6 std m^3/d)	kmol/h = 41.7 (m^3/d)(γ_L)/(MW of liquid)
kg/h = 50 372 (10^6 std m^3/d)(γ_g) = 1739 (10^6 std m^3/d)(MW of gas)	kg/h = 41.7 (m^3/d)(γ_L)
lb–mol/h = 110 (MMscf/d)	lb–mol/h = 14.6 (API bbl/d)(γ_L)/(MW of liquid)
lb/h = 3178 (MMscf/d)(γ_g) = 110 (MMscf/d)(MW of gas)	lb/h = 14.6 (API bbl/d)(γ_L)
Where: MW = molecular weight of gas or liquid	

EFFECT OF PHASE CONDITION

Most of our calculations involve the calculation of Δh ($h_2 - h_1$) or Δs ($s_2 - s_1$) across the one or more pieces of equipment which are being taken as the system for calculation purposes. To obtain these quantities we need to first of all determine the phase condition at the two points involved. Is the stream flowing in or out a subcooled liquid, a saturated liquid, two-phase, a saturated vapor or a superheated vapor?

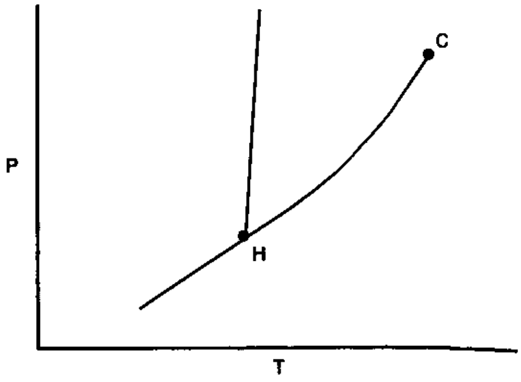
The procedure varies. For a single component system the phase curve is in the general shape of a "Y." Line H-C is the saturated line for liquid and vapor (vapor pressure line). At any pressure and temperature along this line we can have a saturated liquid, two-phases or a saturated vapor. Which of these is present depends on the system energy level.

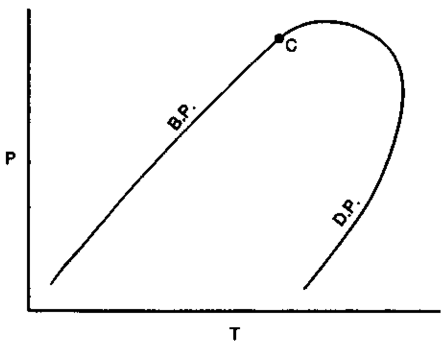
For this single component system we must refer to a correlation of data in figure or table form, or as part of a computer program. These will be discussed in detail in later sections.

For a mixture of two or more components, the phase envelope applies. On the bubblepoint curve the system is a saturated liquid. On the dewpoint curve it is a saturated vapor. Inside the phase envelope, a two-phase system exists.

Mixture computations are less reliable than those for single component systems. It is not convenient to use simple tables or figures. Mixture correlations are based on some mixture physical properties and/or P-V-T equations of state utilizing some form of combination rule.

For a two-phase system, we must find total h or s. This in effect is the sum of the values for vapor and liquid. The academic way to do this is through the use of partial enthalpies and entropies. As a practical matter, Equation 8.4 is easier and just as good.





$$h = (y)(h_V) + (x)(h_L) \tag{8.4}$$

Where:

h = enthalpy of two-phase stream per unit of mass
h_V = enthalpy of saturated vapor per unit of mass
h_L = enthalpy of saturated liquid per unit of mass
y = *wt. fraction* of vapor in total stream (quality)
= $1 - x$
x = *wt. fraction* of liquid in total stream = $1 - y$

This equation can be used also to find total specific volume (υ) and total entropy (s) from saturated values, as shown for h. It says that the total energy or specific volume is the sum of the energy of the two phases present. It may be used for both single and multicomponent systems. The only difference is how one obtains values for the saturated liquid and vapor. In most cases "y" and "x" will be found from a flash calculation when a mixture is involved.

ENTHALPY CALCULATION ALTERNATIVES

There are several basic methods one can use to calculate enthalpy changes between two points in a system.

1. Calculate "sensible heat" changes for a process step where no phase change occurs using Equation 7.15, repeated following as Equation 8.5.
2. For a phase change step calculate "latent heat" from an appropriate correlation.
3. Calculate Δh directly from a correlation for a pure component or a mixture correlation based on overall fluid properties.
4. Calculate Δh from PVT equation of state and basic thermodynamic equations.

Method 4 is the one used for computer solutions. Method 3 also is useful. It is not too often that Method 2 will be used but information is presented for those who choose this approach. Method 1 finds application in high pressure gas (or liquid) heaters and coolers which are essentially isobaric (constant pressure) processes.

SENSIBLE HEAT CHANGES

Equation 8.5 expresses the change in enthalpy for a fluid where no phase change occurs between Points (1) and (2). The second term of this equation is not convenient to solve manually. However, it is trivial or zero for the following cases: (1) ideal gases, (2) constant pressure, $dP = 0$, and (3) for a liquid considered incompressible. For all three cases enthalpy is a mathematical function only of temperature.

C_p is commonly expressed by equations of the form:

$$C_p = A + BT + CT^2$$

Where: A, B, and C are constants that depend on system composition and T is the absolute temperature

In most instances it is sufficiently accurate to find a C_p at the average temperature T_a, where:

$$\Delta h = \int_1^2 C_p dT + \int_1^2 [V - T(\delta V/\delta T)_p]\, dP \tag{8.5}$$

C_{pa} is then found at this average temperature and

$$\int C_p dT = C_{pa}(T_2 - T_1) \tag{8.6}$$

This approximate solution to the first integral, although not exact, is satisfactory for most applications.

HEAT CAPACITY CORRELATIONS

As noted in Chapter 7, values of heat capacity can be found from the slope of h vs. T plots at a given pressure. Enthalpy correlations are shown later in this chapter and in Appendix C at the back of the book. Heat capacity values for pure substances are readily available from many handbooks and similar reference material.

Mixtures

The C_p for hydrocarbon liquid may be estimated from the equation

Metric: $$C_{pa} = 2.96 - 1.34\,\gamma + T\,(0.006\,20 - 0.002\,34\,\gamma) \tag{8.7}$$

English: $$C_{pa} = 0.68 - 0.31\,\gamma + T\,(0.000\,82 - 0.000\,31\,\gamma) \tag{8.7a}$$

Where:		Metric	English
C_{pa} =	heat capacity	kJ/kg·°C	Btu/lb-°F
γ =	liquid sp. gr. (water = 1.0)	–	–
T =	average temperature	°	°

Example 8.1: Find the C_{pa} of a 0.8 sp. gr. oil at an average temperature of 100°C [212°F].

$$C_{pa} = 2.96 - (1.34)(0.8) + 100\,[0.006\,20 - (0.002\,34)(0.8)] = 2.32\ \text{kJ/(kg·°C)}$$

The heat capacity of natural gas may be found from equations of state.(8.1) For most calculations where the use of specific heat is suitable, Figure 8.3 is appropriate. The heat capacity is found at the average temperature and multiplied by gas mass flow rate and ΔT to obtain the heat load.

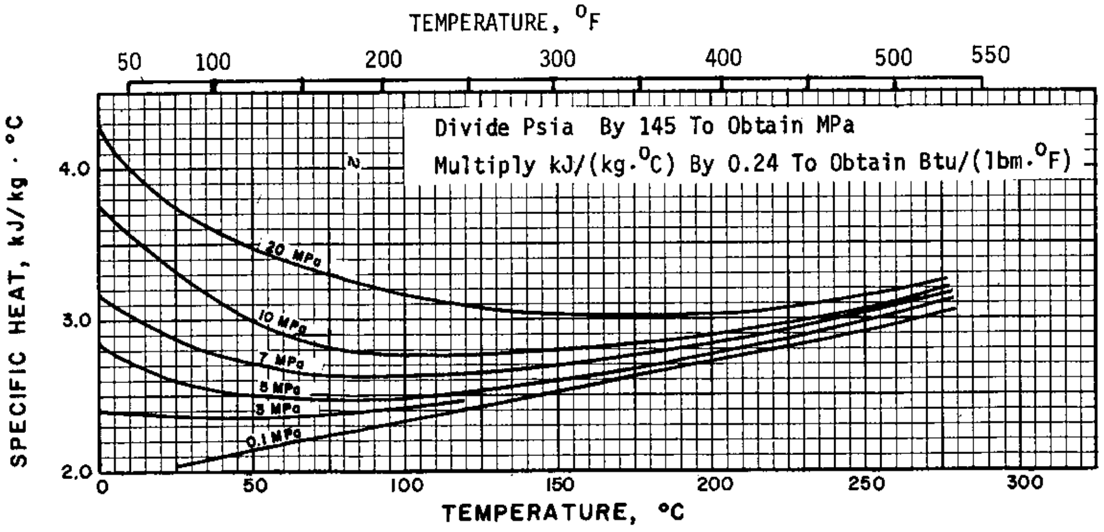


Figure 8.3 **Heat Capacity of 0.65-0.75 Relative Density Natural Gas**

A figure like 8.3 is used primarily for approximate calculations. Normally, the enthalpy change would be calculated directly from a correlation like Appendix 8A or a computer program.

Example 8.2: 10^6 std m^3 per day of a $\gamma = 0.70$ natural gas is being heated from 50 to 100°C in a heater. If the gas pressure is 5.0 MPa, what is the heat load in kJ/h?

From Figure 8.3, Heat Capacity = 2.49 @ average temperature of 75°C

$m = 50\,372\,(1.0)(0.7) = 35\,260$ kg/h

$\Delta H = Q = m\,\Delta h = (35\,260)(2.49)(100 - 50) = 4\,390\,000$ kJ/h $= 1219$ kW

LATENT HEAT

The latent heat is the amount of enthalpy (energy) needed to change phase from liquid to vapor. Energy must be supplied to vaporize; it is released on condensation.

If one uses specific heat in a process involving a phase change, the calculation must be divided into two steps: (1) calculating the sensible heat due to change in P and T, and (2) determining the latent heat change. Total Δh is the sum of these two. If Δh for the process is found by subtracting vapor and liquid enthalpies, phase change effects are included automatically.

Pure Substances

One may calculate latent heat from the table or figure for the substance involved by subtracting the enthalpy values for saturated liquid and saturated vapor.

Sometimes no such table or figure is available. In that case, the following equation may be used.

$$\frac{dP}{dT} = A\left[\frac{\Delta h_L}{T\,(v_g - v_L)}\right] \qquad (8.8)$$

Where:

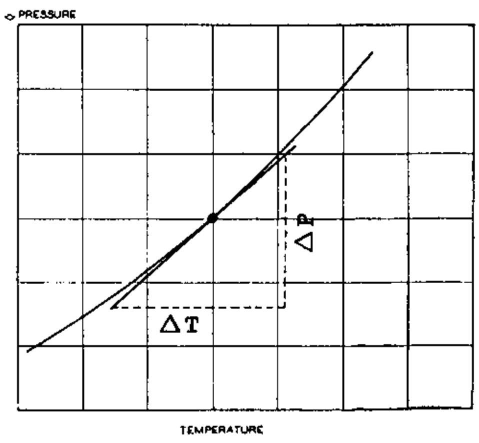
		Metric	English
P	= pressure	kPa	psia
T	= temperature	K	°R
Δh_L	= latent heat	kJ/kg	Btu/lb
v_g	= sp. vol. of vapor	m^3/kg	ft^3/lb
v_L	= sp. vol. of liquid	m^3/kg	ft^3/lb
A	= correlation constant	1.0	5.40

(If the "v" terms are in cm^3/g instead of m^3/kg, A = 1000)

The term v_g is normally much larger than v_L, so v_L can be ignored. If this is done

$$\frac{dP}{dT} = A\left[\frac{(\Delta h_L)(\rho_g)}{T}\right] \qquad (8.9)$$

The value of (dP/dT) may be found from a vapor pressure curve. It is found by obtaining a slope (ΔP/ΔT) in the region of T. The value of ρ_g may be found from the methods given at the beginning of Chapter 3.



Mixtures

Figure 8.4 is a correlation for latent heat based on fluid molecular weight, API gravity and average boiling point. Any two of these three fix the latent heat; use of the first two is recommended when available.

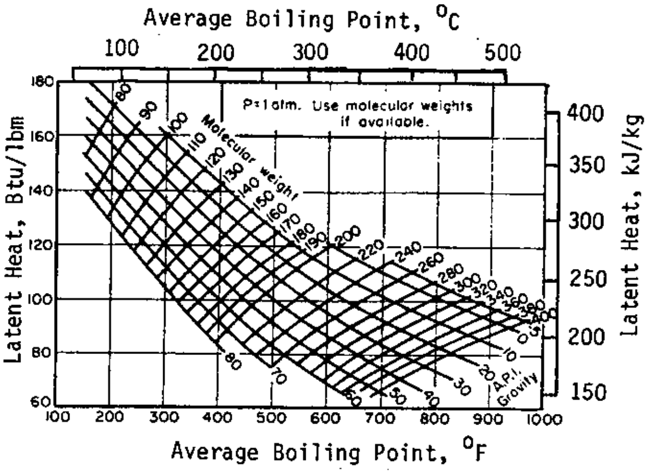


Figure 8.4 **Correlation for Estimating Latent Heat**

The latent heat shown is at atmospheric pressure; it will decrease with increasing pressure. The following is a suitable pressure correction for use with this correlation:

Decrease latent heat 17 kJ/kg for each 1.0 MPa increase above atmospheric pressure [5 Btu/lbm per 100 psi increase].

One can use Figure 8.4 together with Equation 8.7 to find the heat load in a heater where partial vaporization of the liquid occurs.

The latent heat for some mixtures may be found from Appendix 8A. Look up the vapor enthalpy at a given P and T. Subtract the liquid enthalpy at the same T.

The following equation represents another correlation for latent heat of hydrocarbon mixtures at, or near atmospheric pressure.

Metric: $$\Delta h_L = (4.19)(T/M)\,[8.75 + 4.57 \log T] \tag{8.10}$$

English: $$\Delta h_L = (T/M)(7.58 + 4.57 \log T) \tag{8.10a}$$

Where:	Metric	English
T = average boiling point	K	°R
Δh_L = latent heat	kJ/kg	Btu/lb
M = molecular weight		

These equations may be used as an alternative to Figure 8.4.

Example 8.3: A saturated 50° API hydrocarbon liquid is vaporized at 125°F and 700 psia pressure. How much latent heat is involved if the molecular weight is 120?

From Figure 8.4, $\Delta h_L = 130$ Btu/lbm and average boiling point $= 290°F$

From Equation 8.10, $\Delta h_L = (750/120)(7.58 + 4.57 \log 750) = 129$ Btu/lbm

Pressure correction $= (-5)(7) = -35$ Btu/lbm

So, $\Delta h_L = 94 - 95$ Btu/lbm depending on the correlation used.

ENTHALPY CALCULATIONS FOR PURE SUBSTANCES

Figure 8.5 and 8.5(a) are P-H diagrams for propane in metric and English units. All values shown are based on a computer program developed for use with refrigeration processes. The values shown are relative values based on an arbitrary datum at which h and s are taken as zero for a saturated liquid. Other P-H diagrams may show different absolute values for h, however, the Δh will still be correct so long as these relative numbers are consistent.

At any combination of pressure and enthalpy inside the phase curve (saturated liquid and saturated vapor curves) the propane system is two-phase. At a P and h along the saturated liquid curve it is a saturated liquid. It is a saturated vapor along the curve so named.

Propane is a subcooled liquid to the left of the saturated liquid curve. A few lines are shown in Figure 8.5(a) to illustrate that pressure normally has little effect on liquid enthalpy. Lines of constant temperature are horizontal in the two-phase region. At the saturated liquid curve they become essentially vertical except near the critical.

The region to the right of the saturated vapor curve is the superheated vapor region. Knowing any two of the four variables represented fixes a point on the diagram, and thus fixes the value of the remaining two variables.

Figures like these for propane usually check actual values to within ±5-6%, which is typical of such correlations. This error is not signifcant because plant and equipment data are seldom better than this (and often worse).

Example 8.4: What is the latent heat of vaporization of propane at 1 MPa [145 psia]?

Metric: (From Figure 8.5)

$$\Delta h_L = 773 - 437 = \underline{336 \text{ kJ/kg}}$$

English: (From Figure 8.5(a)

$$\Delta h_L = 333 - 188 = \underline{145 \text{ Btu/lb}}$$

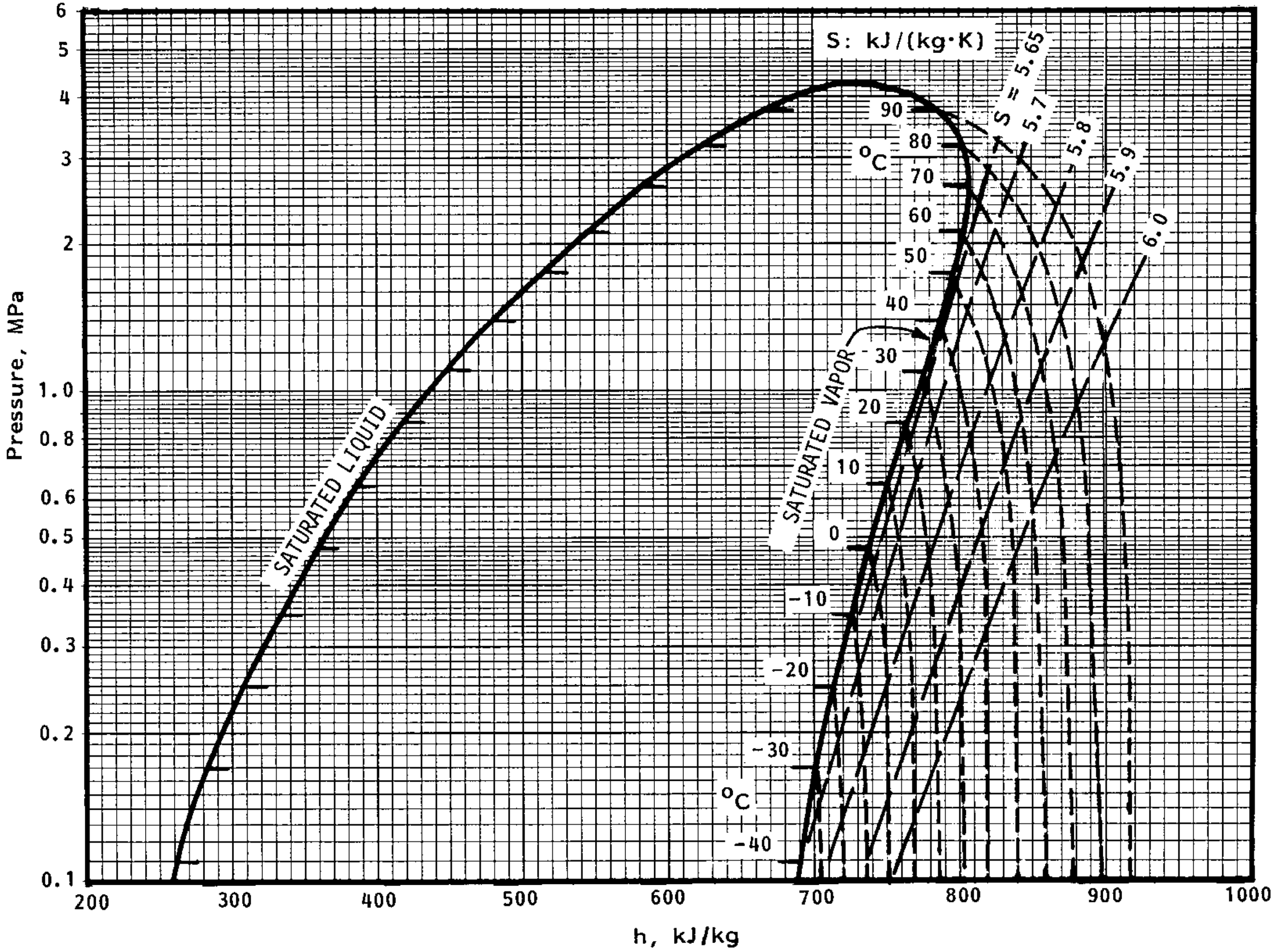


Figure 8.5 **Pressure-Enthalpy Diagram for Propane**

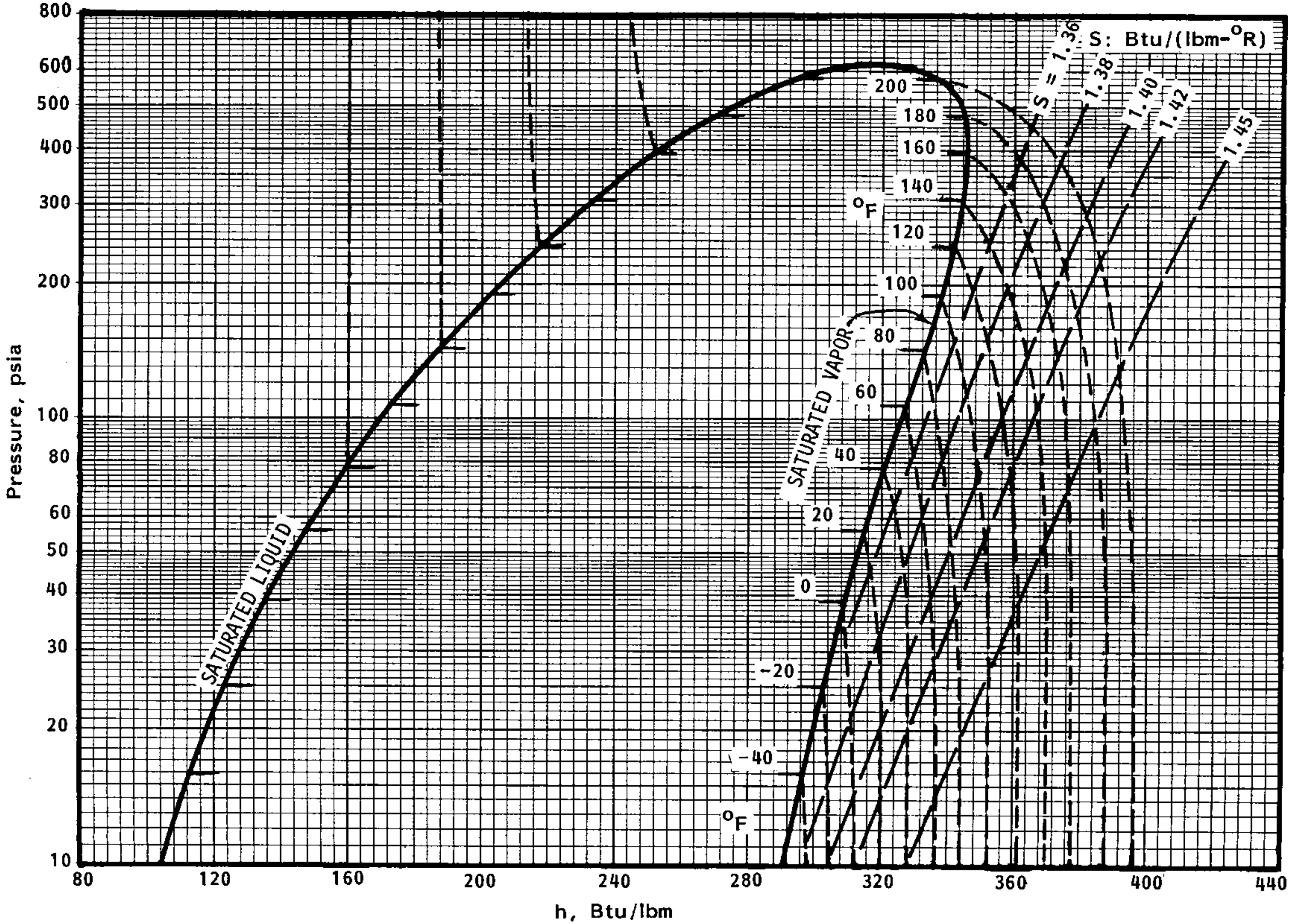


Figure 8.5(a) Pressure-Enthalpy Diagram for Propane

The saturated curves fix the enthalpy of a two-phase mixture at a given pressure by using Equation 8.4. Values for enthalpy are found from the saturated liquid and saturated vapor curves, respectively.

Example 8.5: Calculate the enthalpy of a two-phase propane stream consisting of 60% vapor and 40% liquid at 1 MPa [145 psia].

From Figure 8.5 and 8.5(a)

$$h = (0.6)(773) + (0.4)(437) = 639 \text{kJ/kg}$$

$$\text{or,} \quad h = (0.6)(333) + (0.4)(188) = 275 \text{ Btu/lbm}$$

The same thing can be accomplished with a rule. The total distance between the two curves, at a given pressure, is equivalent to one (x + y). Starting with the saturated liquid curve, measure to the right a distance equivalent to the known "y" (or measure to the left from the saturated vapor curve a distance equivalent to the known "x"). Locate this point and read enthalpy. Or, knowing the mixture enthalpy, locate a point at the known pressure and measure distances x and y.

Table 8.1 and 8.1(a) are for saturated/superheated propane. They are equivalent to Figure 8.5 and 8.5(a). The saturated table applies along the saturation curve of the phase chart (line H-C). The P shown at a given T is the vapor pressure of propane.

Tables of data like this are more precise than figures but the practical accuracy is really no better if one considers the quality of the P and T data normally available.

Consider the previous example for calculation of latent heat of propane at 1 MPa [145 psia]. From the tables (by interpolation) $\Delta h = 333$ kJ/kg or 144 Btu/lbm. Notice that the answers are very close to those from Figure 8.5 but are arrived at by subtracting different numbers. This is another example of the subtraction of *relative energy* values.

Notice in the superheated tables that values of h, v, and s are shown for a series of temperatures at selected pressures. The number in parenthesis behind each selected pressure is the saturation temperature at that pressure. Use arithmetic interpolation to find values between the selected pressures shown.

ENTHALPY CALCULATIONS FOR MIXTURES

Enthalpy may be calculated by one of two basic methods: (1) general correlations and (2) PVT equations of state. The former are used for manual calculations, the latter for most computer solutions. Since enthalpy is dependent on composition, some parameter must be used to characterize the composition of the fluid involved.

General Correlations

A number of useful correlations have been developed over a period of years based on overall fluid properties like molecular weight and density. Correlations of this type have been published in the 1951, 1957, 1966 and 1972 GPSA Engineering Data Books. Other correlations also are available.(8.3-8.5)

Appendix 8A at the end of this chapter is a series of figures for enthalpy as a function of T, P and molecular weight. As illustrated by Figure 8.5(a), the enthalpy of a liquid is essentially independent of pressure. Thus, only one figure is necessary for liquid.

These figures are based on the Mini-Sim version of the SRK equations of state.(8.6) Two sets of figures are enclosed, one in metric and one in English units. Where dashed lines are shown, the actual

TABLE 8.1

THE PROPERTIES OF SATURATED PROPANE

T	P	v_f	v_g	h_f	h_g	s_f	s_g	T	P	v_f	v_g	h_f	h_g	s_f	s_g
–62.22	38.94	1.6543	1011.3	378.14	823.26	3.6819	5.7912	21.11	856.65	2.0033	53.313	571.40	917.21	4.4481	5.6216
–56.67	51.55	1.6731	780.35	389.77	830.23	3.7376	5.7698	26.67	989.66	2.0408	46.509	585.81	921.86	4.4954	5.6158
–51.11	67.40	1.6874	609.92	401.63	837.21	3.7932	5.7527	32.22	1137.2	2.0782	40.141	600.47	926.28	4.5427	5.6103
–45.56	86.84	1.7062	482.57	413.49	843.72	3.8468	5.7368	37.78	1300.5	2.1163	34.835	615.35	930.70	4.5900	5.6053
–40.00	110.27	1.7249	384.56	425.58	850.47	3.9000	5.7234	43.33	1480.4	2.1550	30.402	630.47	934.65	4.6390	5.6011
–34.44	139.08	1.7442	313.39	438.14	857.21	3.9528	5.7108	48.89	1677.5	2.2049	26.594	646.51	939.07	4.6871	5.5969
–28.89	172.64	1.7642	253.46	450.70	863.95	4.0059	5.6982	54.44	1891.8	2.2549	23.098	663.26	942.79	4.7353	5.5919
–23.33	213.30	1.7848	207.88	463.72	870.70	4.0570	5.6865	60.00	2125.4	2.3111	19.977	680.70	946.51	4.7855	5.5881
–17.78	260.58	1.8060	171.05	476.74	877.21	4.1081	5.6752	65.56	2380.4	2.3829	17.355	698.14	949.30	4.8366	5.5793
–12.22	315.99	1.8291	143.58	490.00	883.72	4.1583	5.6652	71.11	2653.3	2.4734	14.983	717.21	950.70	4.8902	5.5697
–6.67	379.05	1.8541	120.49	503.72	889.77	4.2077	5.6564	76.67	2935.9	2.5795	12.985	738.37	950.23	4.9471	5.5567
–1.11	452.79	1.8797	99.885	516.98	895.58	4.2567	5.6484	82.22	3261.2	2.7262	11.237	761.63	947.91	5.0116	5.5362
4.44	536.18	1.9072	83.029	530.00	901.16	4.3053	5.6409	86.67	3432.8	2.8985	9.6763	783.49	943.95	5.0685	5.5136
10.00	630.60	1.9359	71.168	543.72	906.74	4.3534	5.6338	87.78	3607.2	2.9416	9.3017	788.84	940.93	5.0828	5.5082
15.56	736.73	1.9665	61.429	557.21	912.09	4.4007	5.6275	93.33	3962.8	3.2525	7.0543	822.09	926.28	5.1749	5.4596

THE PROPERTIES OF SUPERHEATED PROPANE

T	P = 50.65 kPa (Sat.Temp. = –57.06°C)			P = 84.36 kPa (Sat.Temp. = –46.11°C)			P = 101.28 kPa (Sat.Temp. = –42.06°C)			P = 137.84 kPa (Sat.Temp. = –34.61°C)			P = 206.75 kPa (Sat.Temp. = –24.18°C)			P = 275.67 kPa (Sat.Temp. = –16.17°C)			P = 413.51 kPa (Sat.Temp. = 4.00°C)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
–50.00	822.56	839.39	5.8130																		
–40.00	860.88	853.02	5.8711	510.78	851.63	5.7761	422.95	850.70	5.7401												
–30.00	899.16	866.89	5.9293	533.79	865.81	5.8334	442.48	865.16	5.7988	322.20	863.95	5.7372									
–15.00	955.54	888.34	6.0152	568.29	887.46	5.9194	471.80	887.06	5.8822	343.56	886.33	5.8241	224.94	884.19	5.7428						
0.00	1013.4	911.14	6.1011	602.63	910.37	6.0051	500.80	910.01	5.9900	365.08	908.96	5.9104	239.75	907.14	5.8281	176.96	905.01	5.7679			
10.00	1050.6	927.09	6.1585	625.90	925.96	6.9625	520.19	925.82	5.9681	379.56	925.15	5.9669	249.65	923.21	5.8844	184.58	921.12	5.8257	119.32	916.82	5.7395
20.00	1089.2	942.52	6.2154	648.64	941.65	6.1197	539.28	941.72	6.1909	393.79	940.92	6.0236	259.52	939.23	5.9413	192.25	937.56	5.8826	124.65	933.37	5.8003
30.00	1126.8	961.12	6.2730	672.20	960.62	6.1755	558.49	958.89	6.0170	408.14	957.93	6.0800	269.22	956.20	5.9973	199.64	954.59	5.9388	129.92	950.76	5.8591
40.00	1164.4	973.74	6.3294	694.70	974.31	6.2316	578.11	975.81	6.1281	422.44	974.90	6.1354	278.80	973.74	6.0531	206.95	971.73	5.9956	135.03	968.71	5.9171
50.00	1202.9	994.70	6.3847	717.35	993.61	6.2872	596.88	993.13	6.2527	436.63	992.46	6.1904	288.35	991.26	6.1089	214.25	989.45	6 0520	140.09	986.63	5.9741
60.00	1241.1	1012.1	6.4393	740.39	1013.7	6.3422	616.10	1011.2	6.3070	450.79	1010.7	6.2454	297.91	1009.3	6.1646	221.56	1007.9	6.1077	145.14	1005.1	6.0307
70.00	1278.7	1024.1	6.4936	763.43	1030.3	6.3968	635.36	1029.8	6.3634	464.95	1029.3	6.3005	307.46	1028.1	6.2204	228.86	1026.9	6.1635	150.20	1024.4	6.0872
80.00	1316.8	1046.8	6.5478	786.46	1048.6	6.4511	652.54	1048.9	6.3860	479.11	1048.3	6.3555	317.01	1047.3	6.2762	236.16	1046.3	6.2193	155.26	1044.0	6.1434
90.00	1355.0	1074.5	6.6021	809.52	1069.8	6.5053	669.51	1068.7	6.4262	493.27	1067.7	6.4105	326.56	1067.0	6.3319	243.47	1066.1	6.2746	160.25	1064.1	6.1993

T	P = 551.34 kPa (Sat.Temp. = 5.38°C)			P = 689.18 kPa (Sat.Temp. = 13.12°C)			P = 895.93 kPa (Sat.Temp. = 22.89°C)			P = 1102.7 kPa (Sat.Temp. = 30.95°C)			P = 1309.4 kPa (Sat.Temp. = 38.06°C)			P = 1516.2 kPa (Sat.Temp. = 44.36°C)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
20.00	90.822	928.88	5.7316	70.164	923.71	5.6734												
30.00	95.020	947.11	5.7910	73.857	942.53	5.7417	53.892	934.03	5.6691									
40.00	99.111	964.84	5.8494	77.305	906.77	5.7942	57.207	954.07	5.7322	44.155	945.57	5.6760						
50.00	103.06	982.98	5.9073	80.897	979.58	5.8530	60.159	973.72	5.7939	46.886	966.59	5.7424	37.705	959.75	5.6934	30.924	951.05	5.6438
60.00	107.00	1002.1	5.9654	84.090	999.07	5.9159	62.802	993.72	5.8561	49.368	987.91	5.8058	40.047	981.63	5.7606	33.174	974.88	5.7150
70.00	110.94	1021.8	6.0234	87.050	1018.9	5.9748	65.386	1014.5	5.9176	51.683	1009.0	5.8676	42.173	1003.1	5.8251	35.205	997.00	5.7817
80.00	114.77	1041.6	6.0808	90.373	1039.2	6.0328	67.974	1035.0	5.9780	53.904	1030.4	5.9295	44.171	1025.3	5.8880	37.049	1019.7	5.8470
90.00	118.57	1062.0	6.1375	93.717	1059.9	6.0918	70.523	1055.9	6.0377	56.058	1051.9	5.9908	46.048	1047.2	5.9486	38.786	1042.0	5.9090

T = °C P = kPa
h = kJ/kg v = cm^3/kg,
s = kJ/(kg·k)

TABLE 8.1(a)

PROPERTIES OF SATURATED PROPANE

T	Abs. P, lb/sq in.	Volume, cu ft/lb		Enthalpy, Btu/lb		Entropy, Btu/(lb)(°R)		T	Abs. P, lb/sq in.	Volume, cu ft/lb		Enthalpy, Btu/lb		Entropy, Btu/(lb)(°R)	
	p	Liquid v_f	Vapor v_g	Liquid h_f	Vapor h_g	Liquid s_f	Vapor s_g		p	Liquid v_f	Vapor v_g	Liquid h_f	Vapor h_g	Liquid s_f	Vapor s_g
–80	5.65	0.0265	16.2	162.6	354.0	0.8794	1.3832	70	124.3	0.03209	.854	245.7	394.4	1.0624	1.3427
–70	7.48	0.0268	12.5	167.6	357.0	.8927	1.3781	80	143.6	0.03269	.745	251.9	396.4	1.0737	1.3413
–60	9.78	0.02703	9.77	172.7	360.0	.9060	1.3740	90	165.0	0.03329	.643	258.2	398.3	1.0850	1.3400
–50	12.60	0.02733	7.73	177.8	362.8	.9188	1.3702	100	188.7	0.03390	.558	264.6	400.2	1.0963	1.3388
–40	16.00	0.02763	6.16	183.0	365.7	.9315	1.3670	110	214.8	0.03452	.487	271.1	401.9	1.1080	1.3378
–30	20.18	0.02794	5.02	188.4	368.6	.9441	1.3640	120	243.4	0.03532	.426	278.0	403.8	1.1195	1.3368
–20	25.05	0.02826	4.06	193.8	371.5	.9568	1.3610	130	274.5	0.03612	.370	285.2	405.4	1.1310	1.3356
–10	30.95	0.02859	3.33	199.4	374.4	.9690	1.3582	140	308.4	0.03702	.320	292.7	407.0	1.1430	1.3347
0	37.81	0.02893	2.74	205.0	377.2	.9812	1.3555	150	345.4	0.03817	.278	300.2	408.2	1.1552	1.3326
10	45.85	0.02930	2.30	210.7	380.0	.9932	1.3531	160	385.0	0.03962	.240	308.4	408.8	1.1680	1.3303
20	55.00	0.02970	1.93	216.6	382.6	1.0050	1.3510	170	426.0	0.04132	.208	317.5	408.6	1.1816	1.3272
30	65.70	0.03011	1.60	222.3	385.1	1.0167	1.3491	180	473.2	0.04367	.180	327.5	407.6	1.1970	1.3223
40	77.80	0.03055	1.33	227.9	387.5	1.0283	1.3473	190	523.4	0.04712	.149	339.2	404.6	1.2140	1.3156
50	91.50	0.03101	1.14	233.8	389.9	1.0398	1.3456	200	575.0	0.0521	.113	353.5	398.3	1.2360	1.3040
60	106.9	0.03150	0.984	239.6	392.2	1.0511	1.3441								

PROPERTIES OF SUPERHEATED PROPANE

T	7.35 (–70.7°F)			12.24 (–51°F)			14.696 (–43.708°F)			20 (–30.30°F)			30 (–11.52°F)			40 (2.90°F)			60 (24.80°F)			80 (41.69°F)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	12.72	356.5	1.3785	7.90	362.75	1.3718	6.66	364.6	1.3681	5.050	368.4	1.3640	3.30	374.00	1.3588	2.61	378.2	1.3548	1.76	383.8	1.3501	1.32	387.8	1.3470
–60	13.11	360.3	1.3869																					
–40	13.79	366.8	1.4023	8.182	366.2	1.3796	6.775	365.8	1.3710															
–20	14.47	373.4	1.4177	8.591	373.0	1.3948	7.123	372.8	1.3866	5.186	372.2	1.3719												
0	15.14	380.2	1.4329	9.000	379.9	1.4100	7.471	379.8	1.4020	5.439	379.3	1.3873	3.559	378.3	1.3678									
20	15.82	387.4	1.4481	9.409	387.0	1.4252	7.816	386.8	1.4772	5.695	386.5	1.4025	3.735	385.7	1.3830	2.753	384.7	1.3684						
40	16.50	394.8	1.4633	9.818	394.4	1.4404	8.160	394.3	1.4324	5.951	393.9	1.4177	3.911	393.1	1.3980	2.889	392.2	1.3838	1.863	390.1	1.3630			
60	17.17	402.4	1.4785	10.23	402.0	1.4556	8.503	401.9	1.4474	6.207	401.6	1.4327	4.087	400.8	1.4130	3.025	400.0	1.3990	1.959	398.2	1.3789	1.424	396.1	1.3624
80	17.85	410.2	1.4937	10.64	410.0	1.4706	8.844	409.8	1.4623	6.461	409.4	1.4477	4.261	408.7	1.4280	3.159	408.0	1.4140	2.053	406.3	1.3948	1.500	404.6	1.3785
100	18.52	418.4	1.5088	11.050	418.1	1.4854	9.187	418.0	1.4772	6.716	417.6	1.4625	4.432	417.0	1.4428	3.289	416.2	1.4290	2.145	414.8	1.4102	1.573	413.2	1.3940
120	19.20	426.6	1.5235	11.45	426.2	1.5002	9.528	426.2	1.4918	6.969	425.9	1.4771	4.602	425.4	1.4576	3.419	424.6	1.4440	2.235	423.4	1.4254	1.644	421.8	1.4094
140	19.88	435.2	1.5380	11.86	435.9	1.5148	9.869	434.8	1.5064	7.221	434.6	1.4917	4.772	434.0	1.4724	3.549	433.4	1.4588	2.325	432.2	1.4404	1.714	430.9	1.4248
160	20.55	444.1	1.5524	12.27	443.8	1.5293	10.21	443.7	1.5210	7.473	443.5	1.5063	4.942	443.0	1.4872	3.679	442.5	1.4736	2.415	441.4	1.4554	1.784	440.3	1.4402
180	21.23	453.1	1.5668	12.68	452.9	1.5437	10.51	452.9	1.5256	7.725	452.6	1.5209	5.112	452.2	1.5020	3.809	451.8	1.4884	2.505	450.8	1.4703	1.852	449.8	1.4554
200	21.90	462.5	1.5812	13.09	462.4	1.5581	10.84	462.4	1.5500	7.977	461.9	1.5355	5.282	461.8	1.5168	3.939	461.3	1.5030	2.593	460.5	1.4851	1.920	459.7	1.4704

T	100 (55.62°F)			130 (73.20°F)			160 (87.71°F)			190 (100.50°F)			220 (111.85°F)			250 (122.12°F)			300 (137.55°F)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	1.06	391.2	1.3448	0.8165	395.0	1.3424	0.6685	397.9	1.3404	0.5540	400.4	1.3388	0.4738	402.5	1.3375	0.4130	404.2	1.3355	0.3332	406.6	1.3345
60	1.094	393.5	1.3488																		
80	1.164	402.6	1.3656	0.8456	398.8	1.3486															
100	1.227	411.4	1.3816	0.9045	408.3	1.3659	0.6969	404.7	1.3521												
120	1.289	420.3	1.3962	0.9588	417.8	1.3822	0.7464	414.6	1.3698	0.5995	411.6	1.3580	0.4911	407.7	1.3460						
140	1.347	429.6	1.4130	1.006	427.3	1.3987	0.7908	424.8	1.3867	0.6415	422.1	1.3759	0.5314	419.2	1.3650	0.4473	415.9	1.3550	0.3392	408.7	1.3372
160	1.400	439.1	1.4286	1.052	437.2	1.4150	0.8319	434.9	1.4031	0.6792	432.4	1.3930	0.5673	429.8	1.3827	0.4816	427.1	1.3732	0.3745	422.0	1.3580
180	1.460	448.8	1.4440	1.098	447.0	1.4310	0.8712	445.1	1.4195	0.7144	443.0	1.4096	0.5998	440.6	1.3999	0.5121	438.0	1.3908	0.4037	433.2	1.3765
200	1.516	458.8	1.4598	1.143	457.1	1.4468	0.9093	455.4	1.4357	0.7472	453.4	1.4255	0.6302	451.2	1.4161	0.5408	449.9	1.4074	0.4303	444.9	1.3944

T = °F P = psia
h = Btu/lb p = lb/in^2
s = Btu/(lb)(°R)
v = ft^3/lb

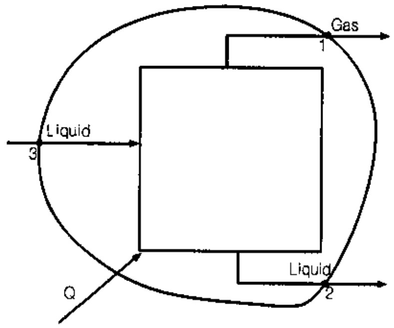
Paranthetic figures after pressures are saturation temperatures.

hydrocarbon system may not be in the phase shown in the figure. That portion of the curve is, in effect, an extrapolation of calculated data.

Values are shown for liquid molecular weights from 30-100. Above this range other correlations are advised. One alternative would be the use of Equations 8.7 and 8.10. Below this range one is looking at methane-ethane liquid for which this correlation is not suitable.

Values are shown for gas molecular weights from 16-30. This is considered the practical range for this type of correlation. Pressures of 100 kPa-14.0 MPa [15-2000 psia] are plotted. To find a gas enthalpy at a pressure between those shown, simply use arithmetic interpolation.

The figures in Appendix 8A are useful for early planning, troubleshooting and general manual calculations. It is simple to do and the results generally are satisfactory. The figures are compatible with the computer programs developed to go with this book for typical hydrocarbon streams composed primarily of paraffin hydrocarbons.



Suppose one has a process in which vaporization or condensation is occurring as shown at left. One simply looks up the h at each of the points shown in order to complete the balance. For this case, $Q = m_1h_1 + m_2h_2 - m_3h_3$. The enthalpy of gas is higher than that of liquid to account for latent heat.

The enthalpy at Points (2) and (3) could be found from a liquid enthalpy correlation such as Table 8.1. The vapor enthalpy at Point (1) would be found from a gas enthalpy correlation at the pressure of the process.

This could be an oil heater where a portion of the oil vaporizes. There will be a small pressure drop involved but this will have a negligible effect on the enthalpies. An average pressure could be used with no loss in practical accuracy for the value of Q obtained.

The small pressure change in most heat exchangers has little effect on the energy balance. At constant temperature, enthalpy increases with decreasing pressure but a very small change in pressure is not statistically significant with most enthalpy correlations.

In solving a vaporization and condensation problem (or any problem involving multiple streams) remember that mass in equals mass out and energy in equals energy out, if there is no mass accumulation in the system in question. For our system above

$$m_1 + m_2 = m_3 \qquad \text{and} \qquad m_1h_1 + m_2h_2 = m_3h_3 + Q$$

Once the mass balance has been completed and values of "h" found, the heat load "Q" can be calculated.

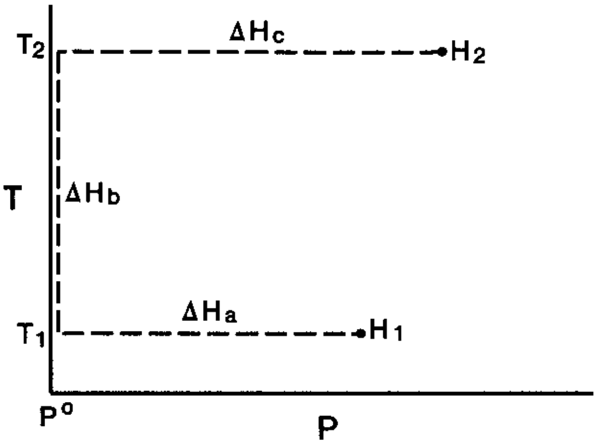
PVT Equations of State

Enthalpy is a point function. It depends only on initial and final conditions and not on the path taken by a process between those points. Thus, one might as well pick a very convenient path for calculation purposes.

Shown below is the path normally chosen for computer solutions along with the appropriate general equations.

The pressure P^o is designated as any pressure at or near atmospheric at which ideal gas correlations are suitable. All temperature changes (Step b) are taken at this pressure. All pressure changes are taken in isothermal steps as shown. The equations written to express these isobaric enthalpy changes often are called *enthalpy departure* equations.

A host of PVT equations has been used in this procedure. Many are about equally applicable.(8.2) The real question is not which equation (by name) was used but the extent and quality of data used to obtain the empirical constants in that equation. Is the correlation being used on the same kinds of systems as those from which it was developed? Does this correlation apply at the P and T conditions involved? Oftentimes the user does not possess these answers. So, he is using the program "blindly." Unfortunately, use of anything blindly, including computer programs, is fraught with peril. Remember: you – not the computer – are responsible to your employer for the quality of your output.



$$\Delta H = H_2 - H_1 = \Delta H_a + \Delta H_b + \Delta H_c$$

$$\Delta H_a = H_1^o - H_1$$

$$\Delta H_b = \int C_p dT$$

$$\Delta H_c = H_2 - H_1^o$$

No computer program is best for all systems. If in doubt, determine enthalpy changes by several independent methods to obtain a range of reasonable answers. Use these results in a judgmental manner to specify equipment.

ENTROPY AND INTERNAL ENERGY CALCULATIONS

It is seldom, if ever, that one will calculate specific internal energy values. However, if the need ever arises remember that

$$U = H - PV$$

As you will see in Volume 2, entropy is involved in the calculations for pumps, compressors and expanders. Although specific values for entropy seldom are tabulated, they are an integral part of the calculation.

It is common also to calculate entropy from enthalpy. As shown in Equation 7.14.

$$dH = TdS + VdP$$

At constant pressure $dP = 0$ and $\Delta S = \Delta H/T$. To prove this, look in Table 8.1 for saturated propane. Divide the Δh at any P and T by the absolute T. The result is the value of ΔS shown (subject to roundoff).

This principle offers a convenient way to calculate entropy changes.

To find an entropy consistent with the enthalpy values shown in Appendix 8A, simply divide the enthalpy value by the appropriate absolute temperature, °R or K, at the pressure at that point in the system.

REFERENCES

8.1 Abou-Kassem, J. H. and Dranchuk, P. M., SPE Paper No. 10980, Presented at Fall Meeting (1982).

8.2 Maddox, R. N. and Erbar, J. H., *Gas Conditioning and Processing*, Vol. 3, Campbell Petr. Series, Norman, OK (1982).

8.3 Canjar, L. N. and Peterka, V. J., *AIChE Jour.*, Vol. 2 (1956), p. 343.

8.4 Kesler, M. G. and Lee, B. I., *Hydr. Proc.* (March 1976), p. 153.

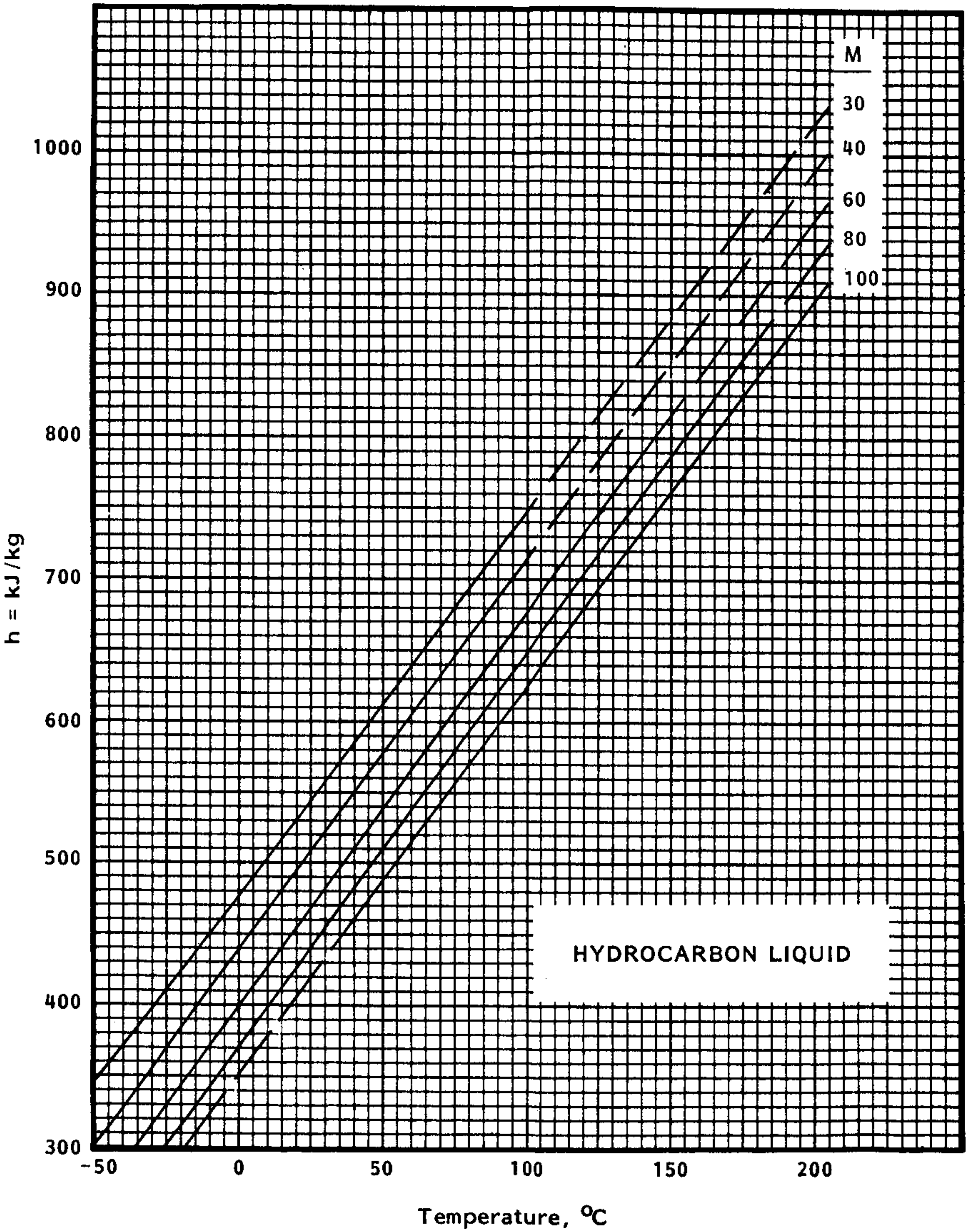
8.5 Katinas, J. G. and Danner, R. P., *Ibid.* (March 1977), p. 157.

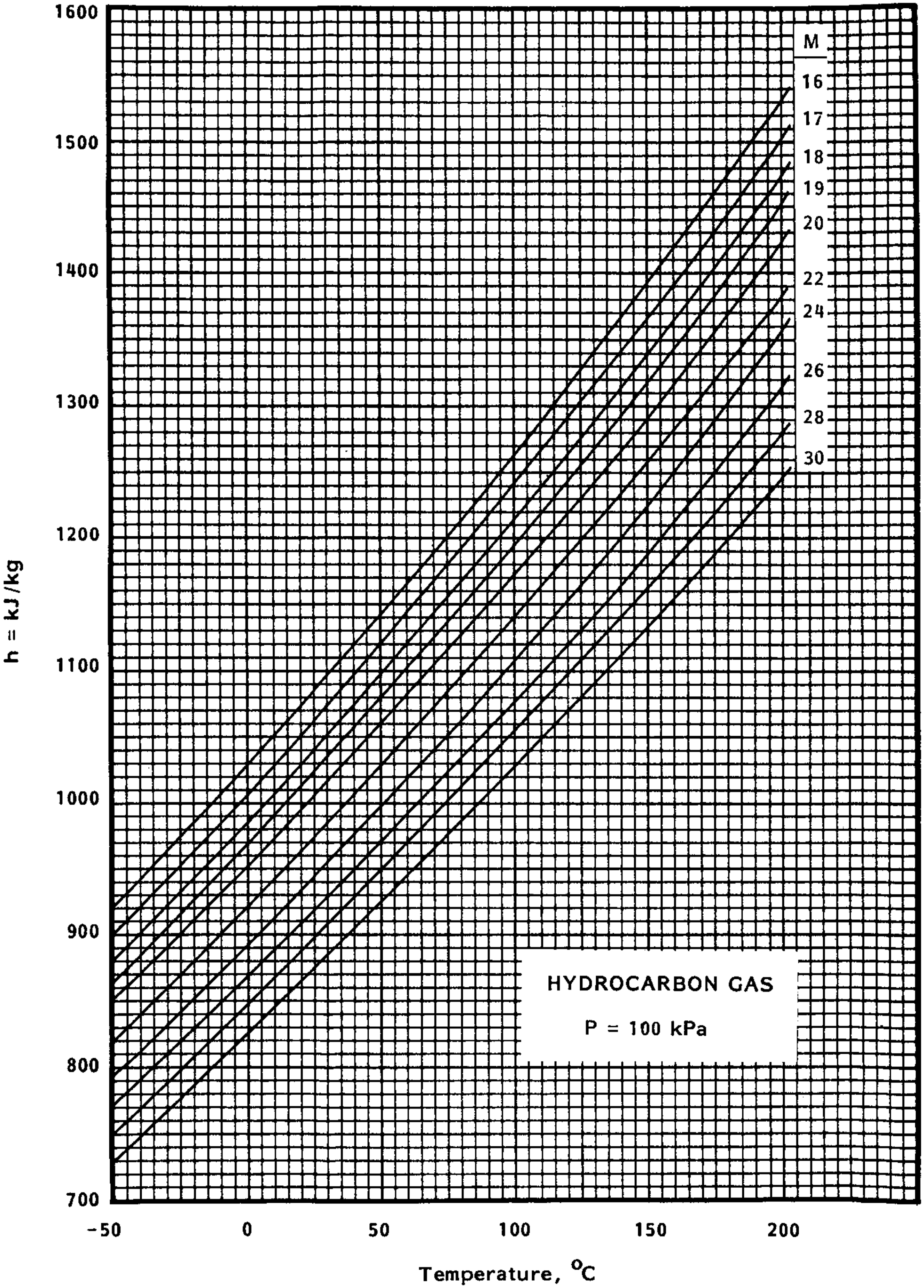
8.6 Maddox, R. N., private communication.

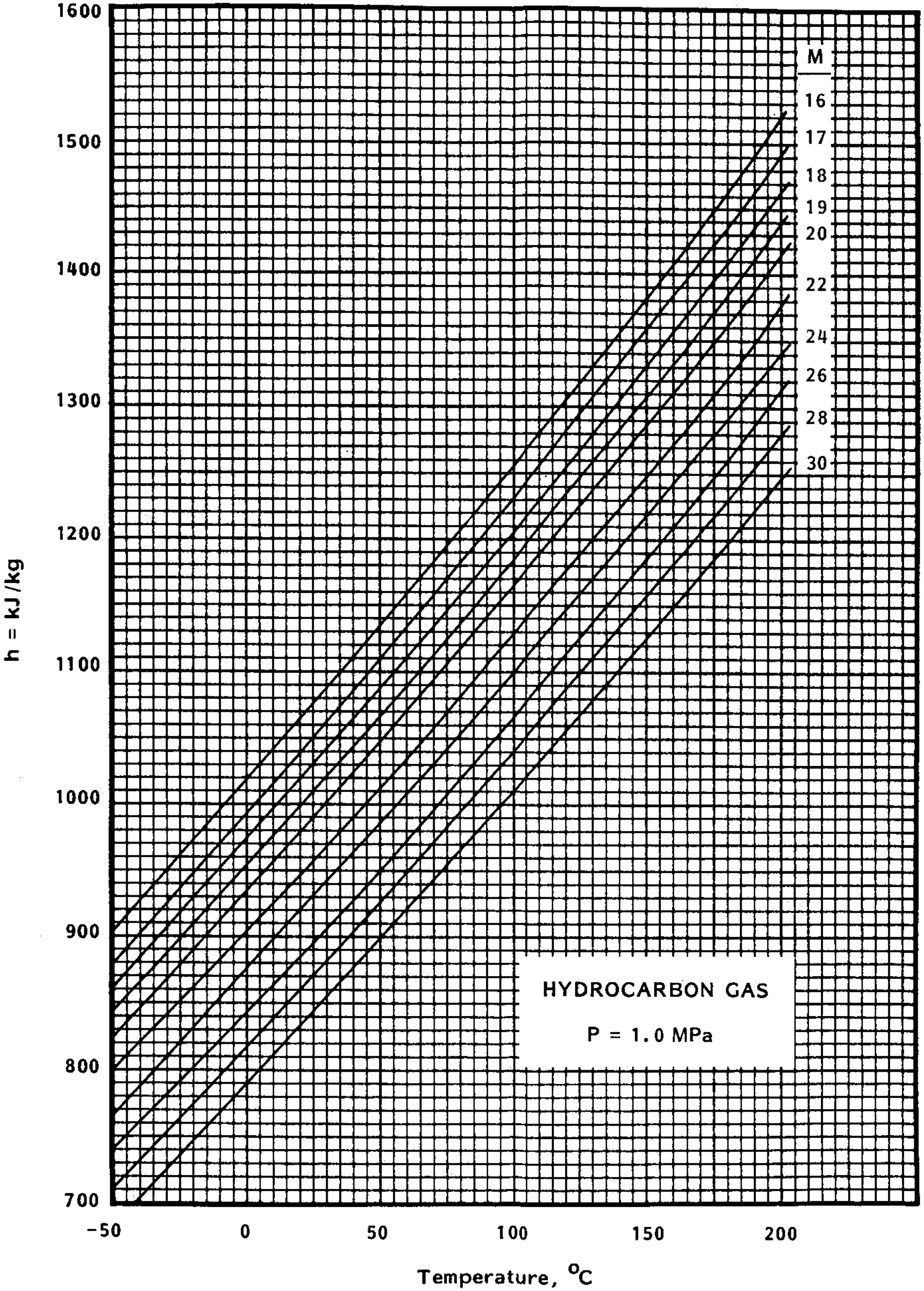
APPENDIX 8A

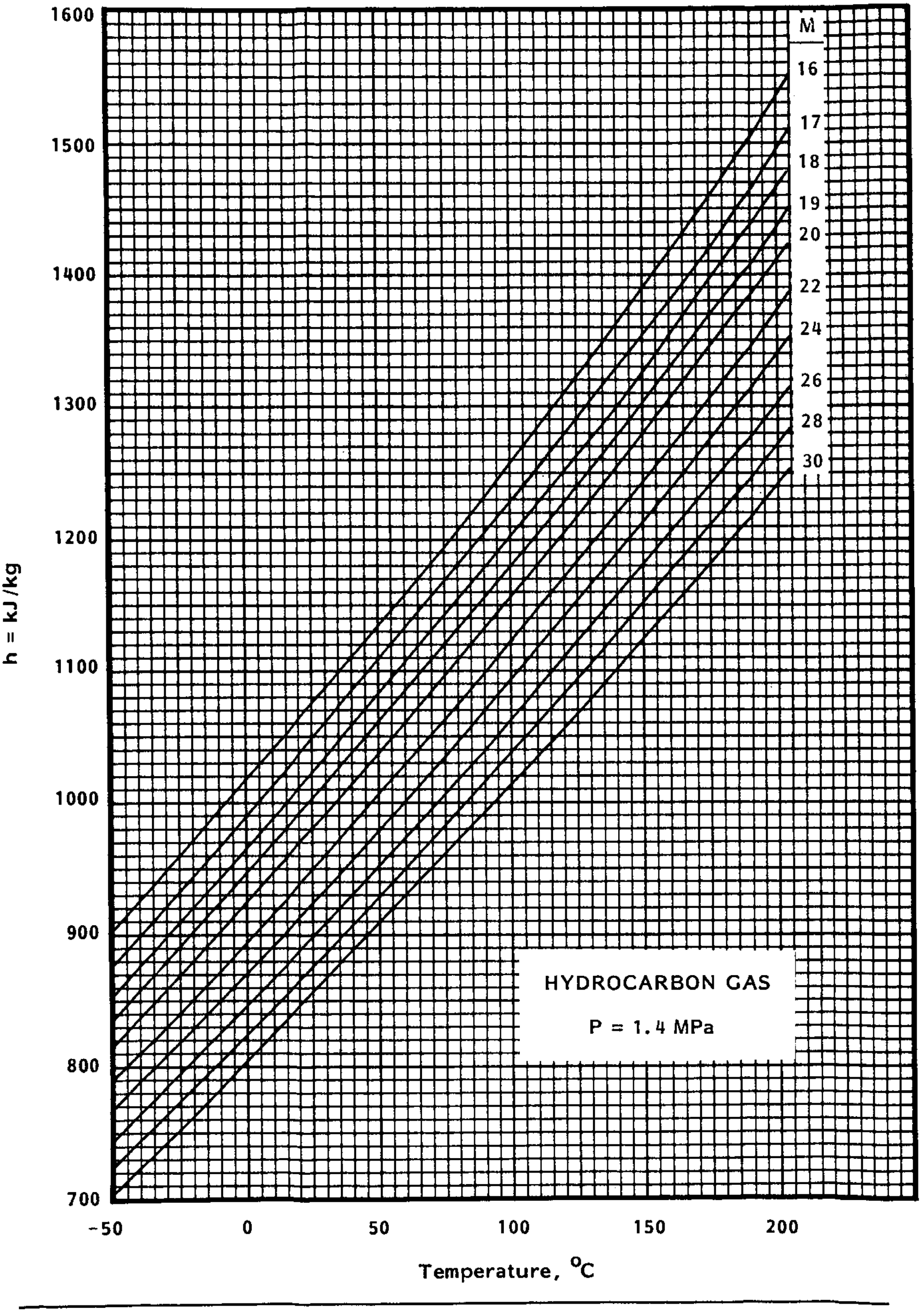
There are two sets of figures, one in Metric and one in English units. Where lines are dashed the correlation is less exact than where solid lines are shown.

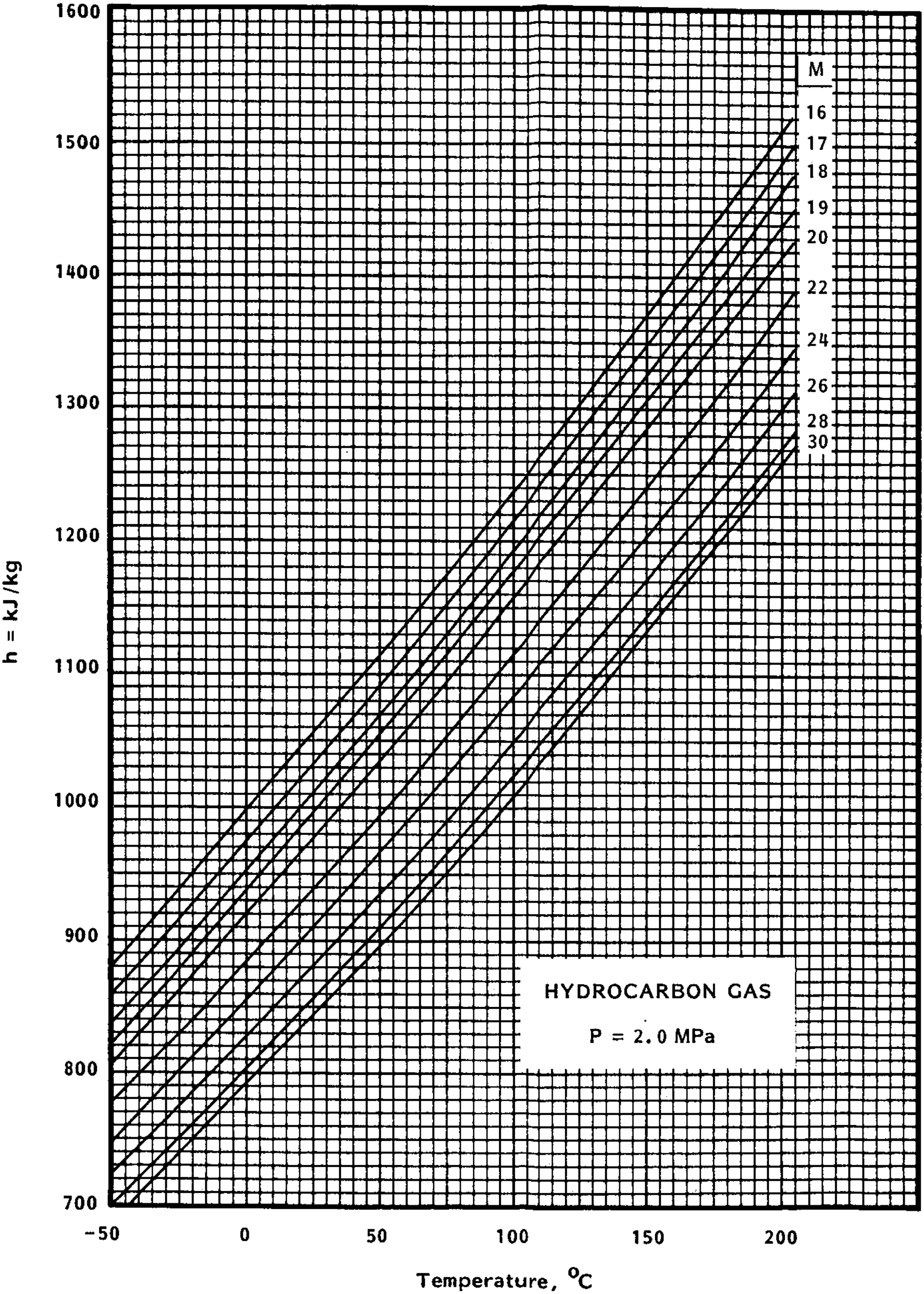
These figures are useful for estimations of enthalpy necessary for planning, specification and operations. Since enthalpy is composition sensitive they are not recommended for detailed design calculations.

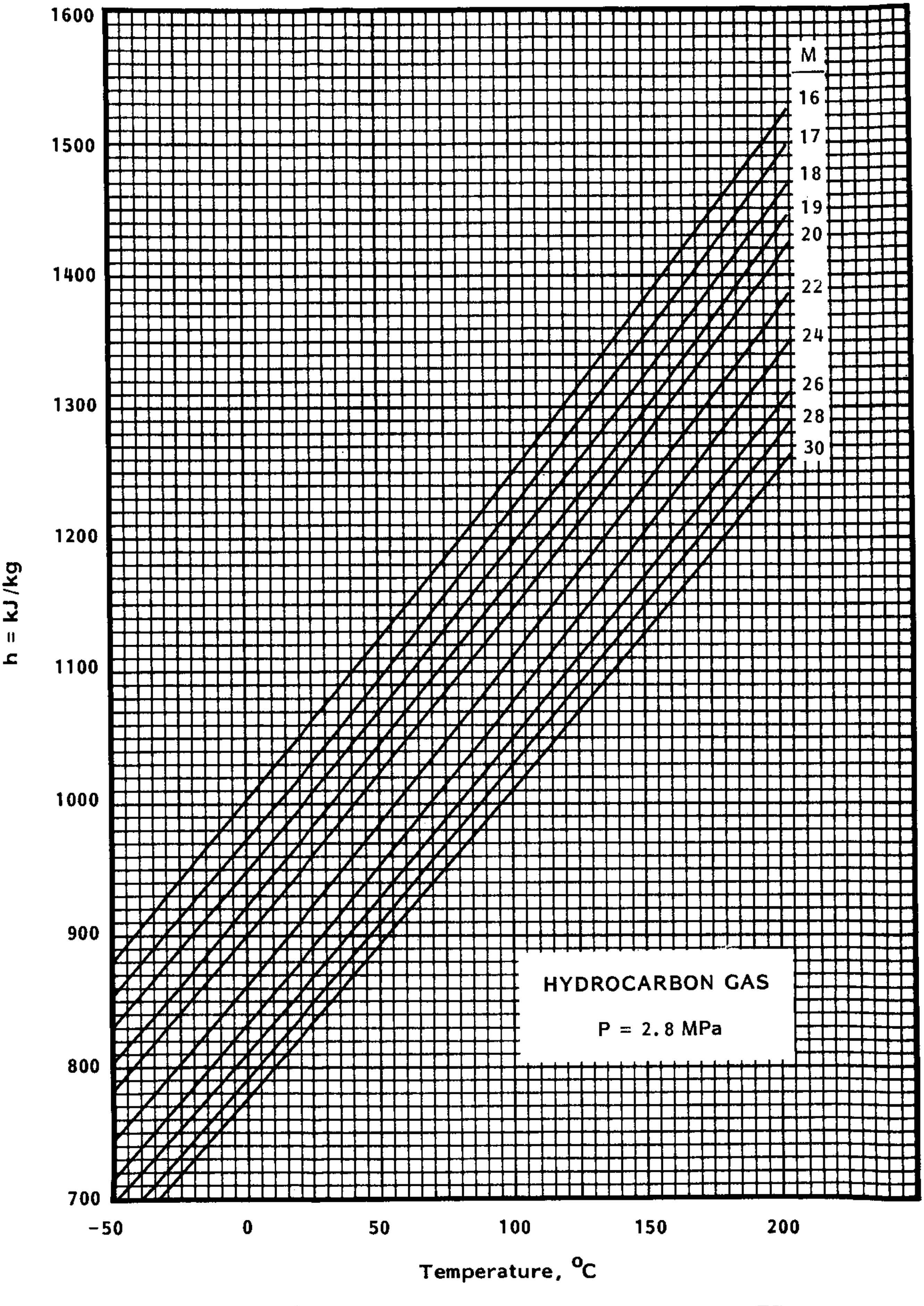

M
30
40
60
80
100
1000
900
800
700
600
500
400
300
h = kJ/kg
-50
0
50
100
150
200
Temperature, °C
HYDROCARBON LIQUID

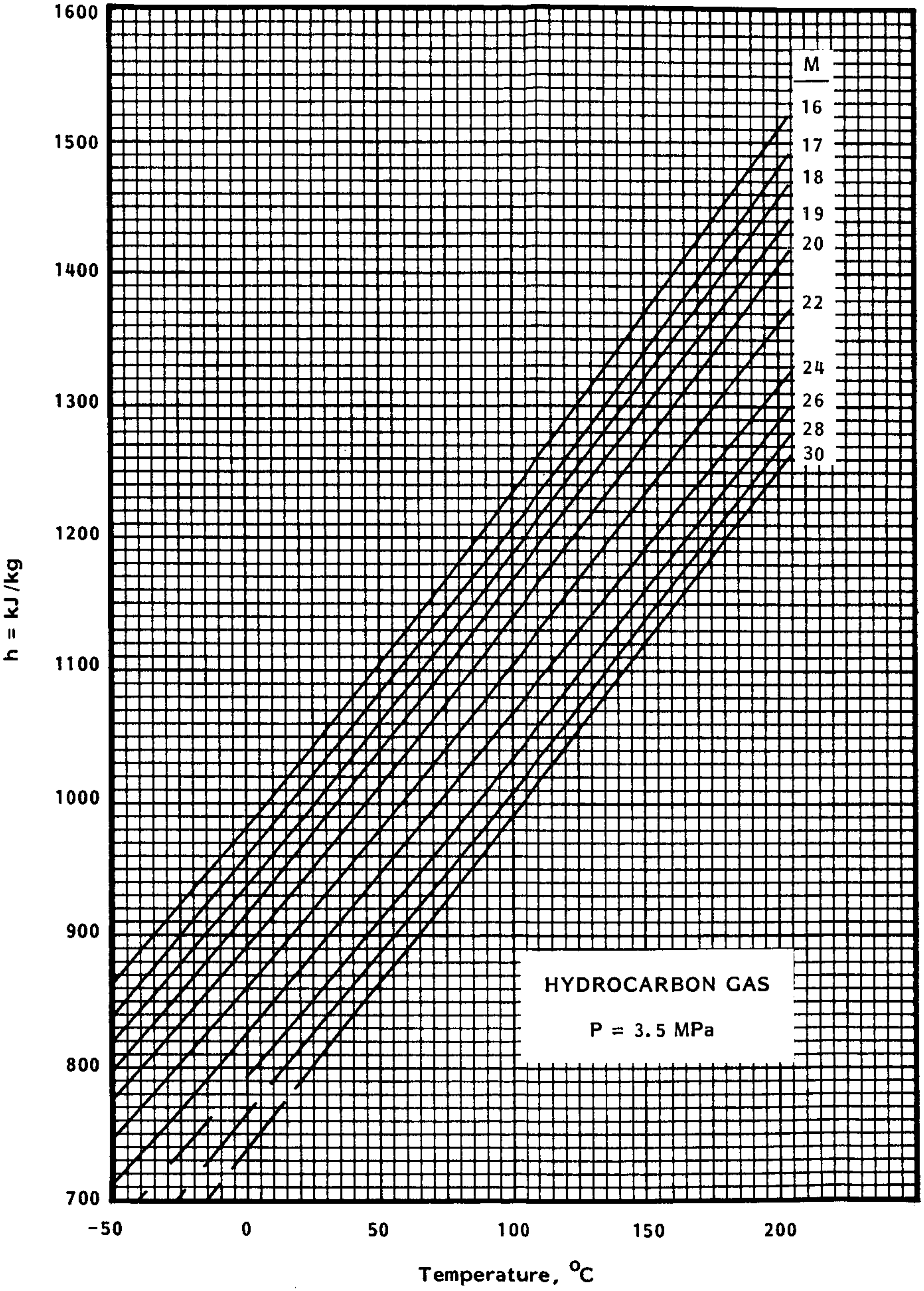

h = kJ/kg
1600
1500
1400
1300
1200
1100
1000
900
800
700
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 100 kPa

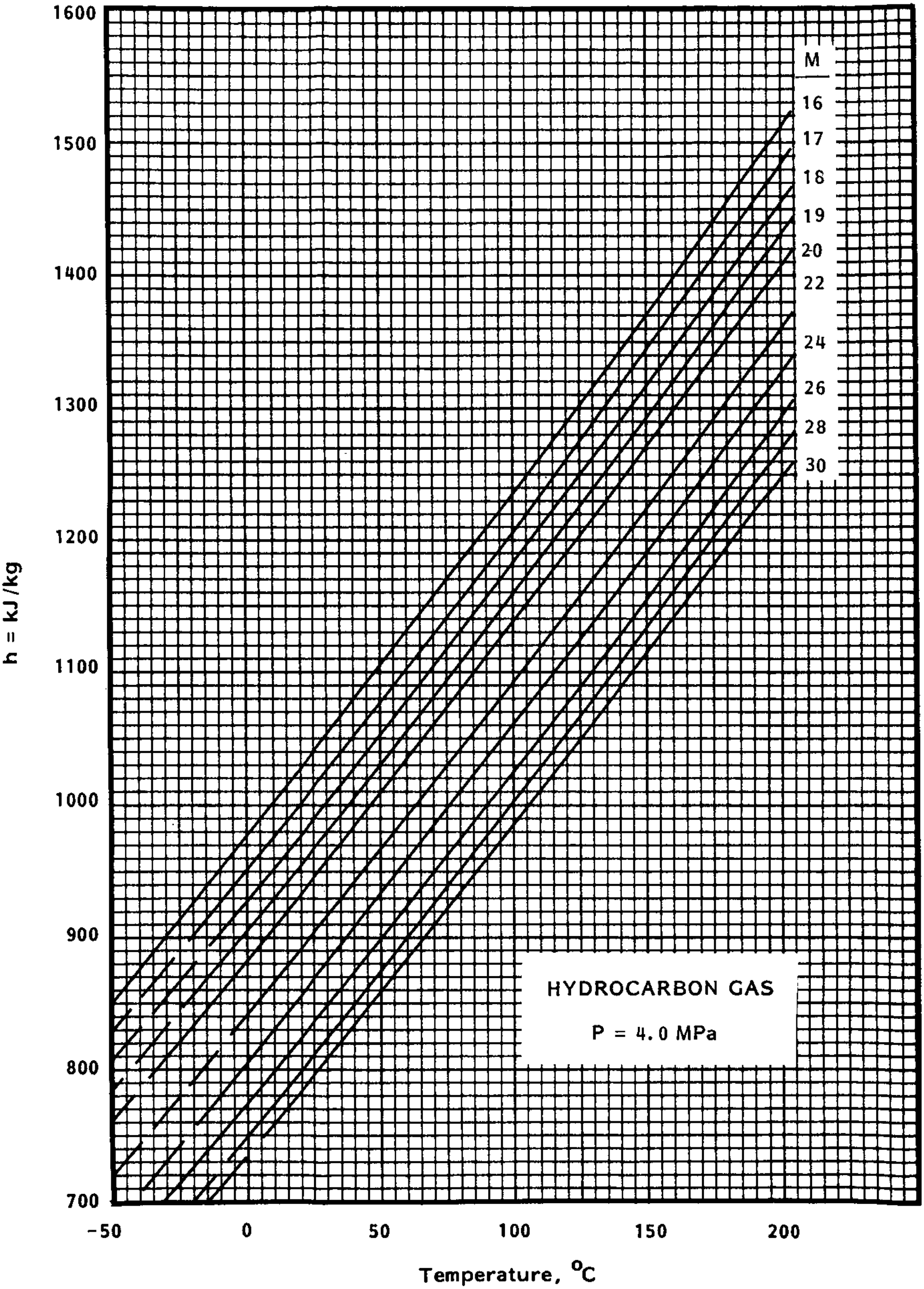

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 1.0 MPa
−50
0
50
100
150
200
Temperature, °C

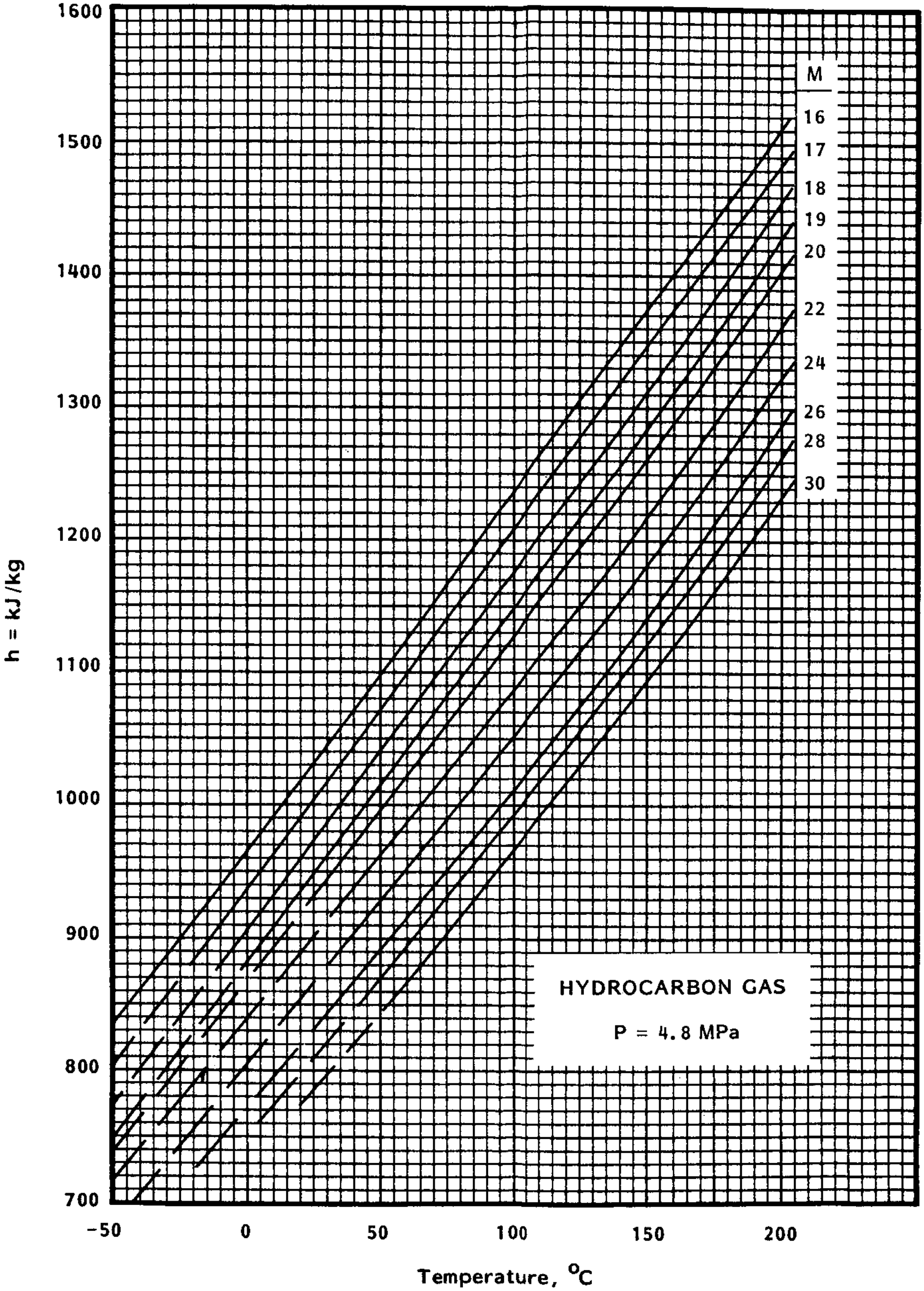

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 1.4 MPa

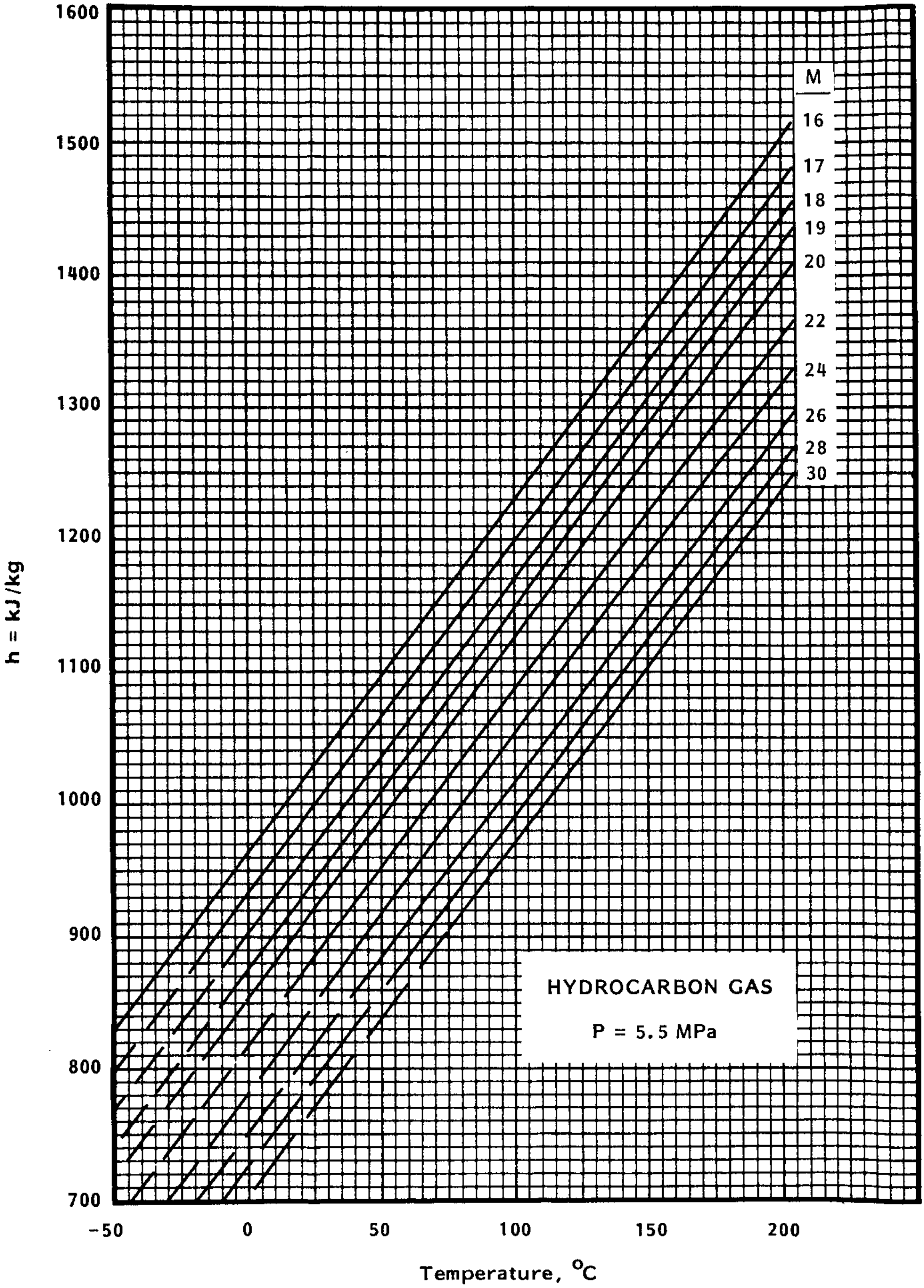

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
−50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 2.0 MPa

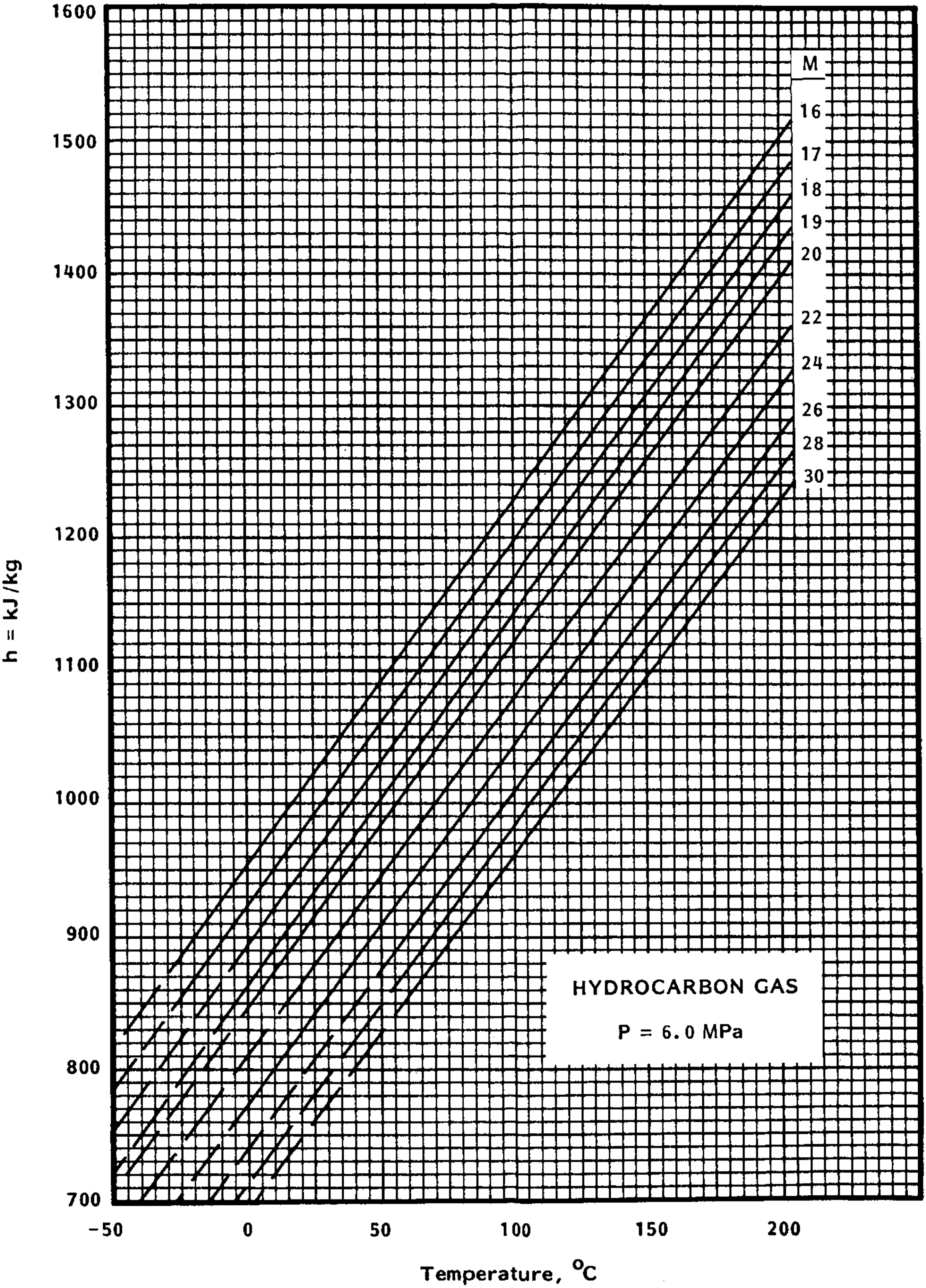

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 2.8 MPa

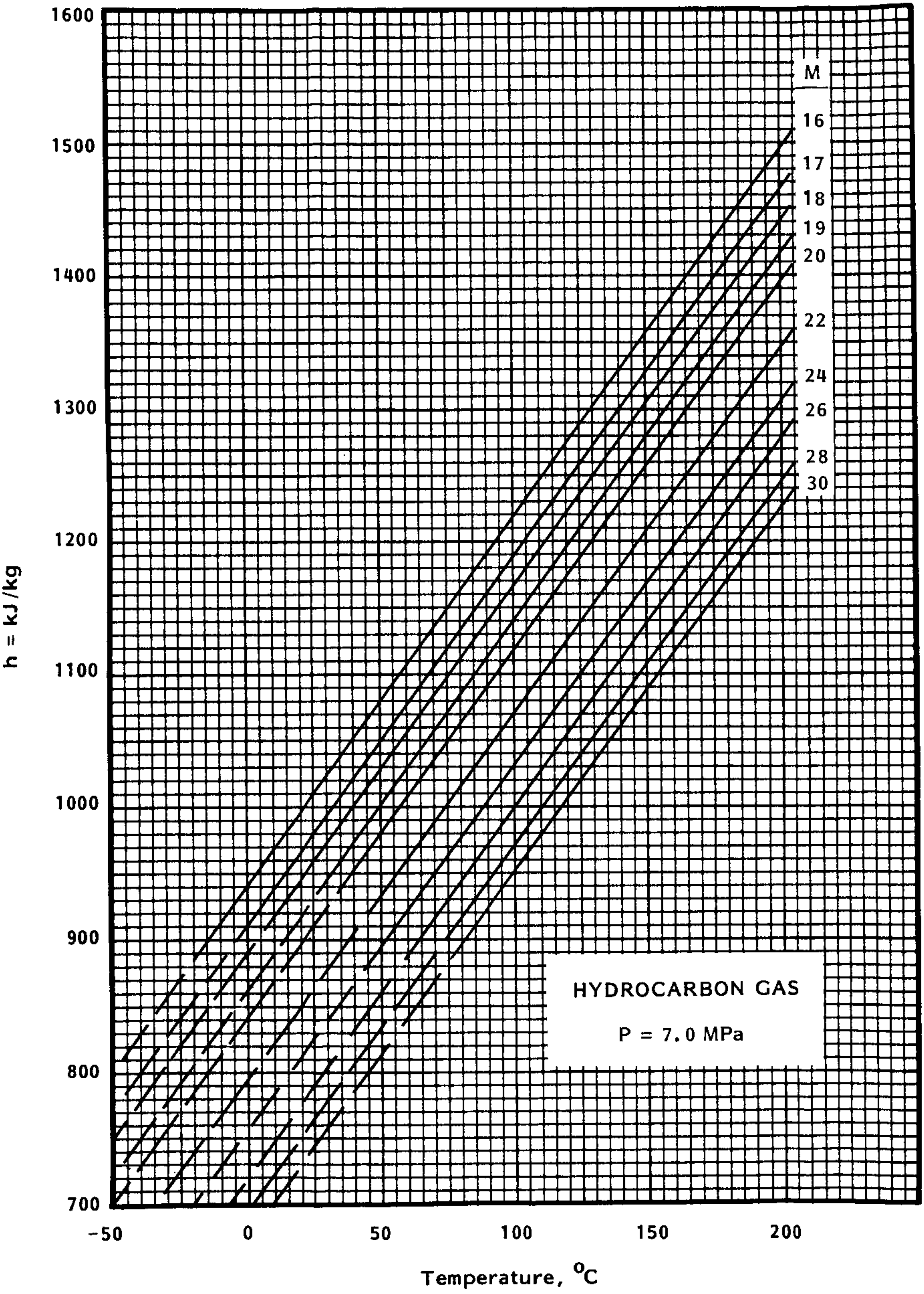

M
16
17
18
19
20
22
24
26
28
30
1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
-50
0
50
100
150
200
Temperature, °C
HYDROCARBON GAS
P = 3.5 MPa

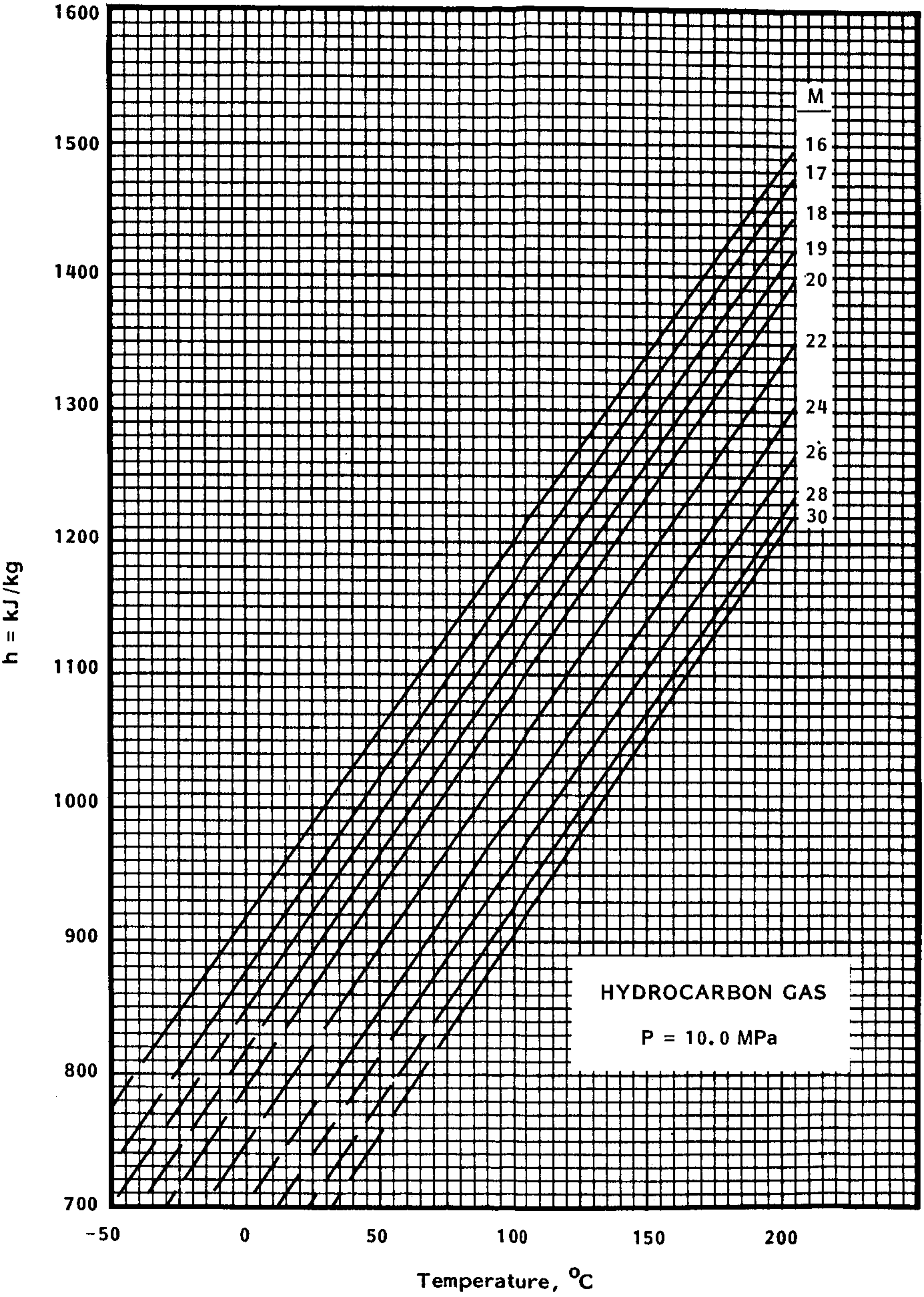

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ /kg
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 4.0 MPa

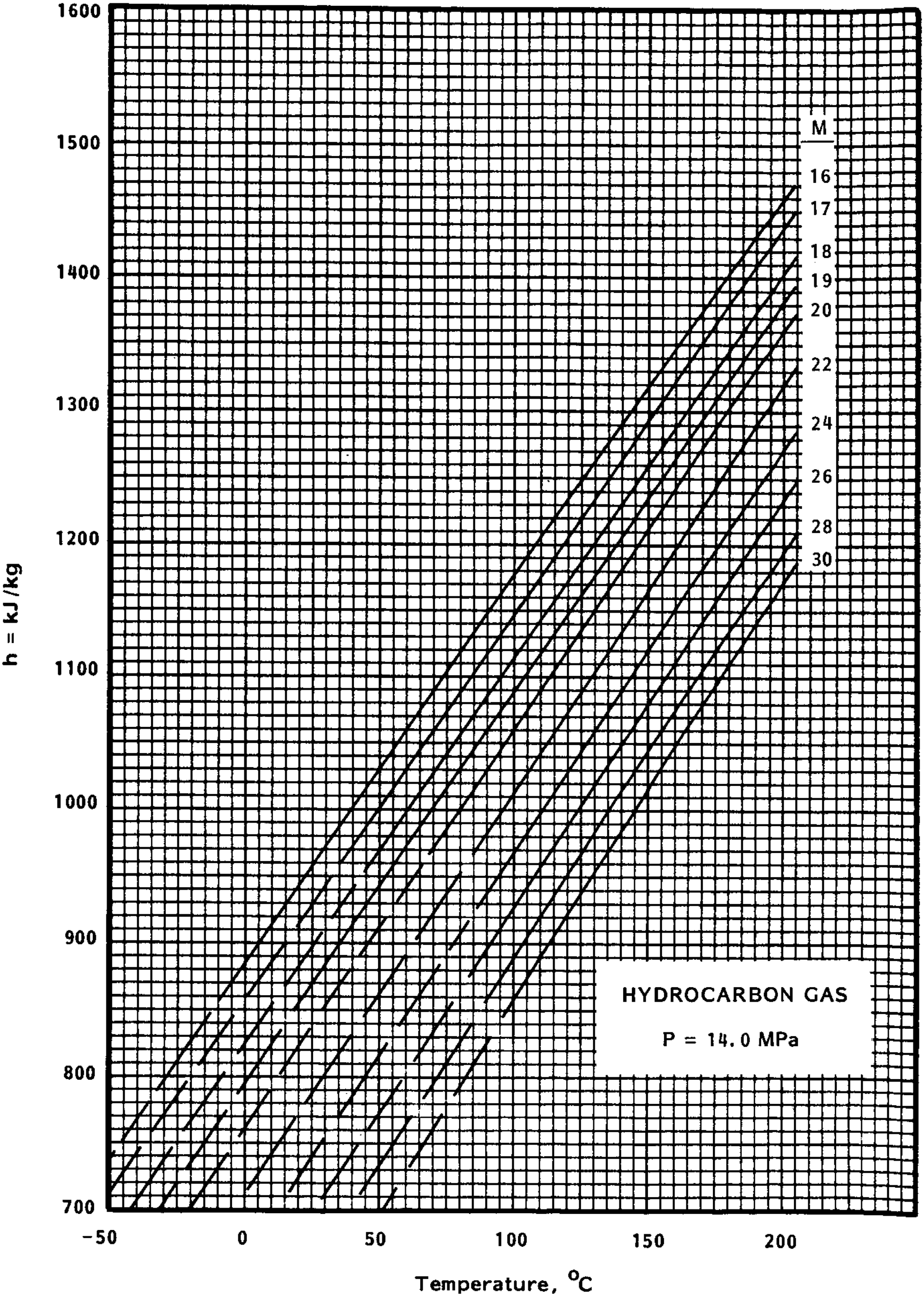

h = kJ/kg
1600
1500
1400
1300
1200
1100
1000
900
800
700
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 4.8 MPa

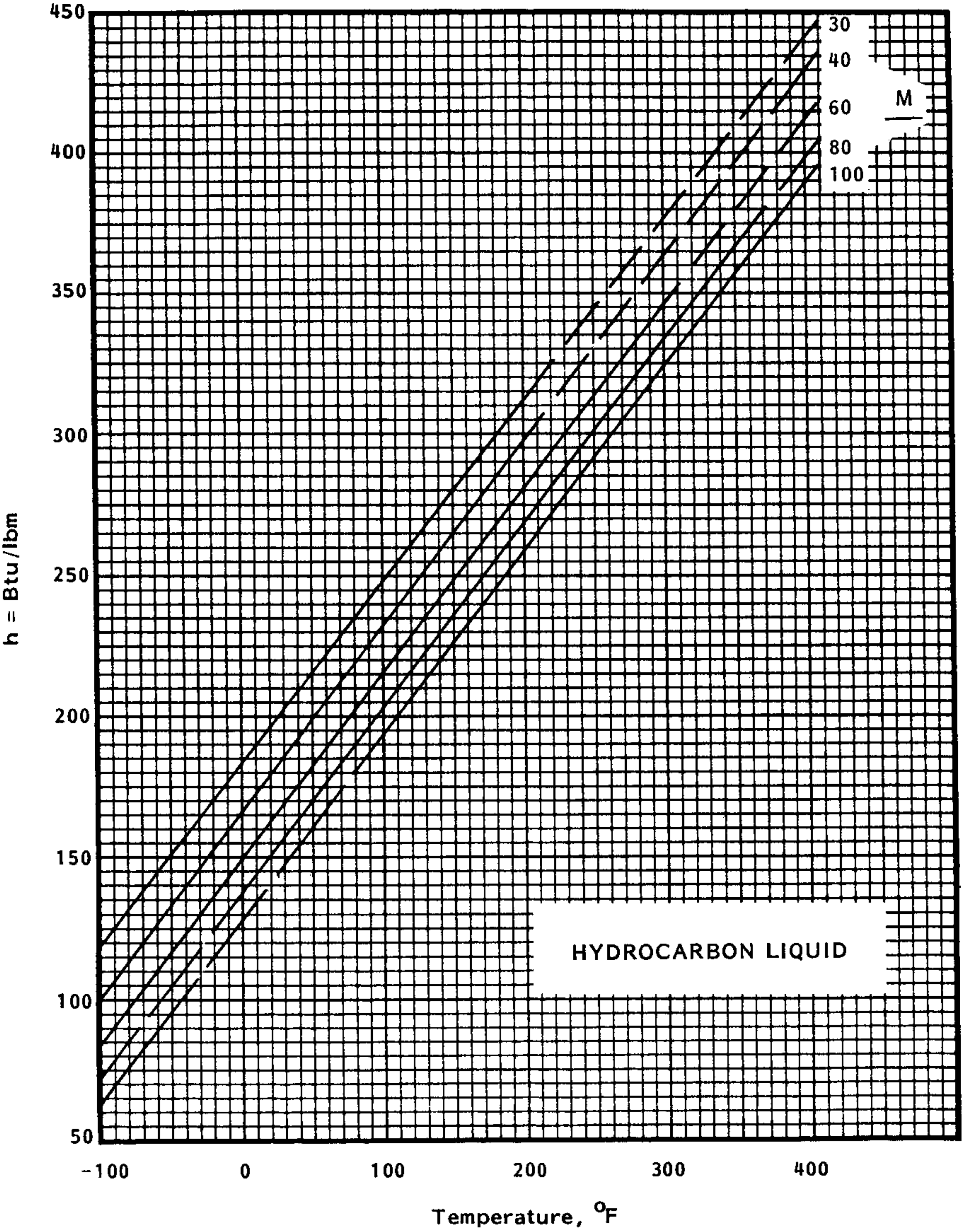

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 5.5 MPa
-50
0
50
100
150
200
Temperature, °C

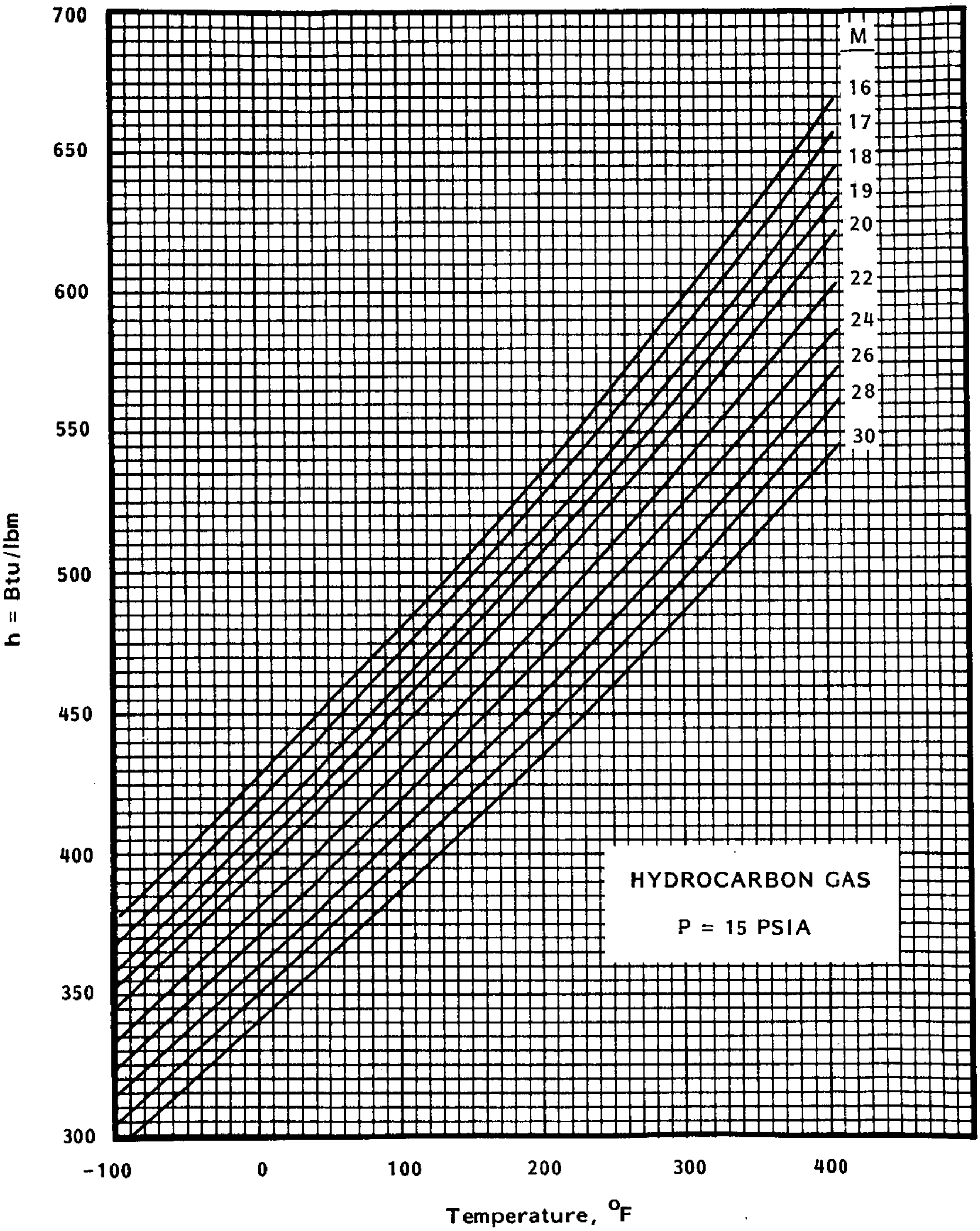

1600
1500
1400
1300
1200
1100
1000
900
800
700
h = kJ/kg
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 6.0 MPa

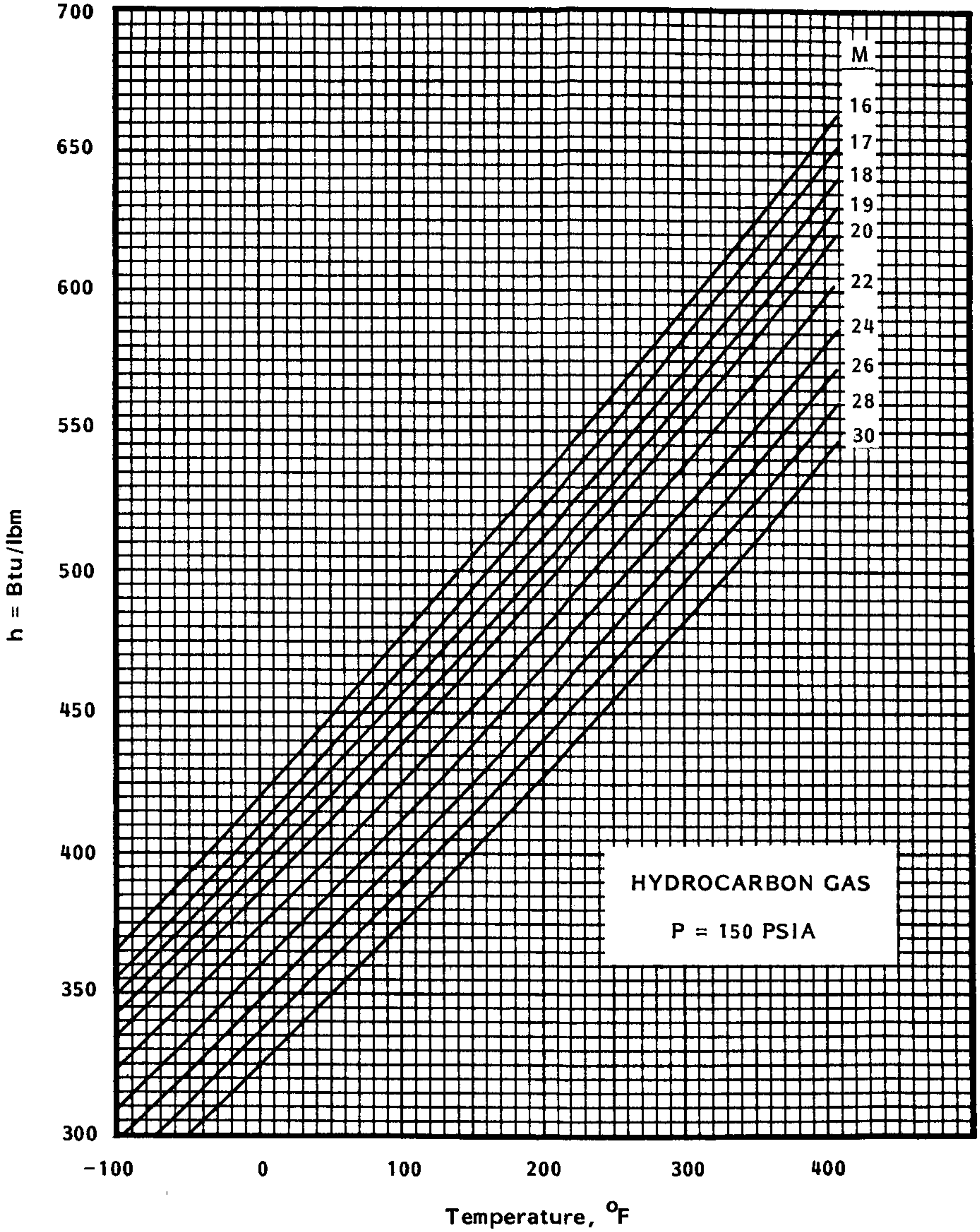

h = kJ/kg
1600
1500
1400
1300
1200
1100
1000
900
800
700
-50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 7.0 MPa

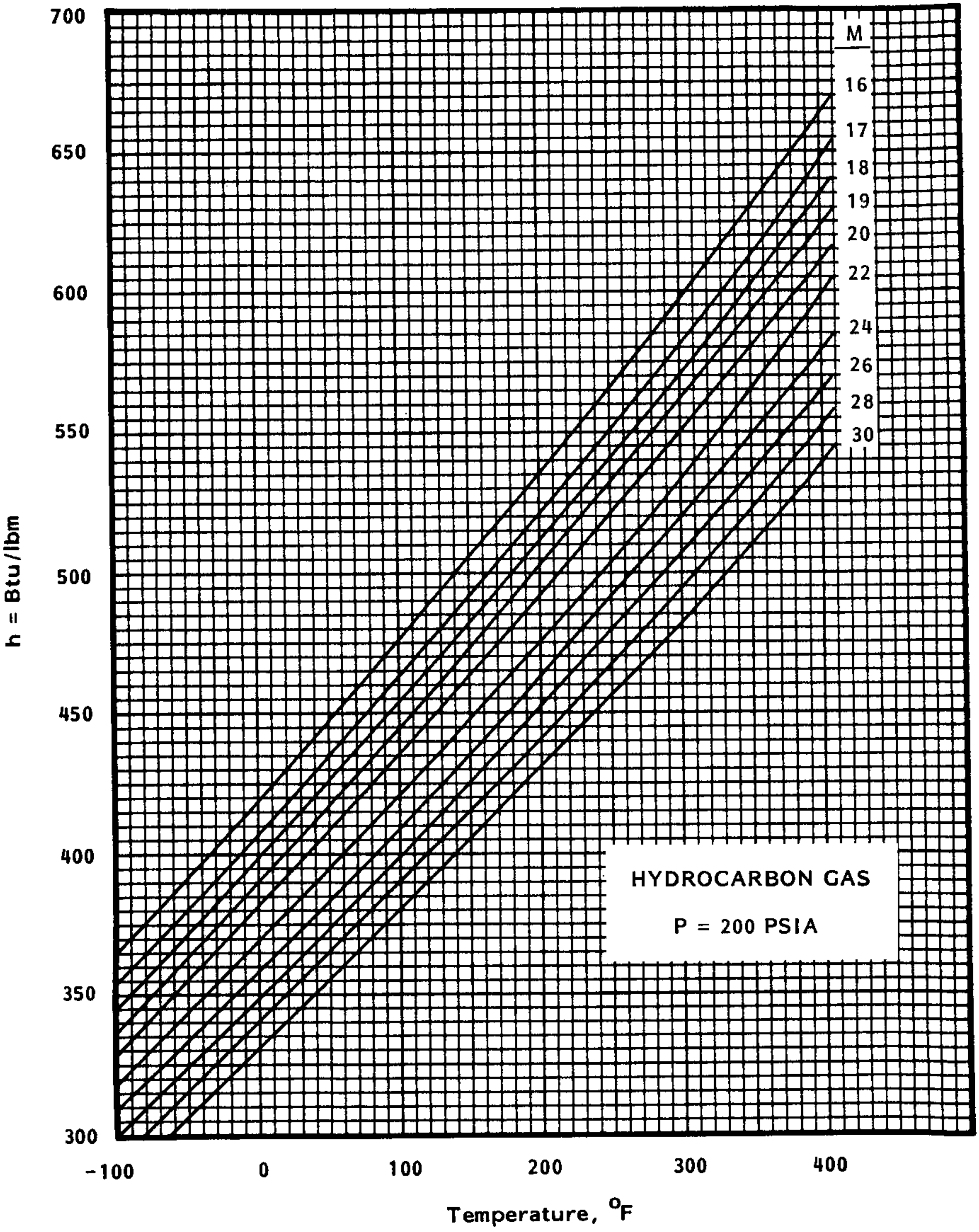

h = kJ/kg
1600
1500
1400
1300
1200
1100
1000
900
800
700
−50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 10.0 MPa

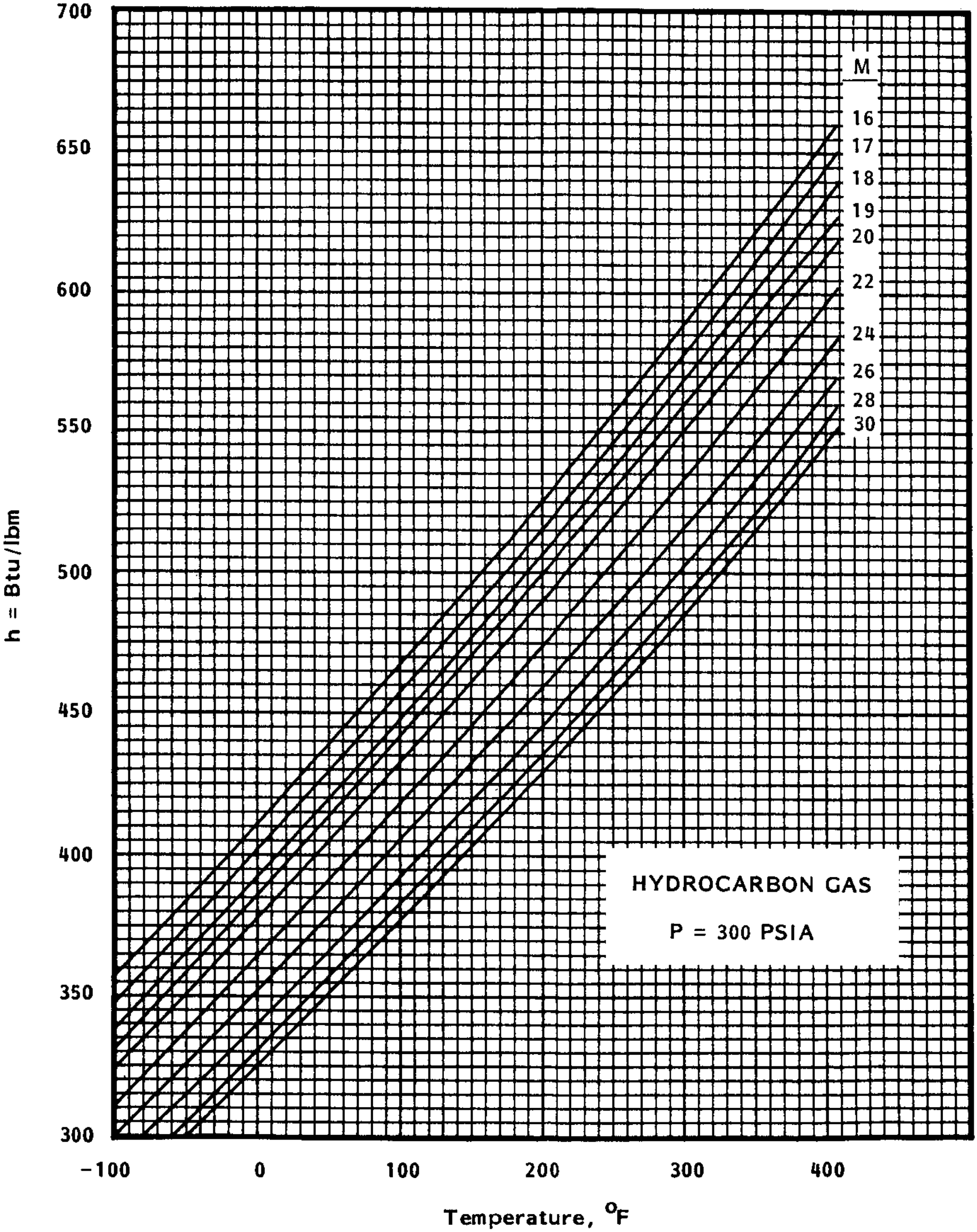

h = kJ/kg
1600
1500
1400
1300
1200
1100
1000
900
800
700
−50
0
50
100
150
200
Temperature, °C
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 14.0 MPa

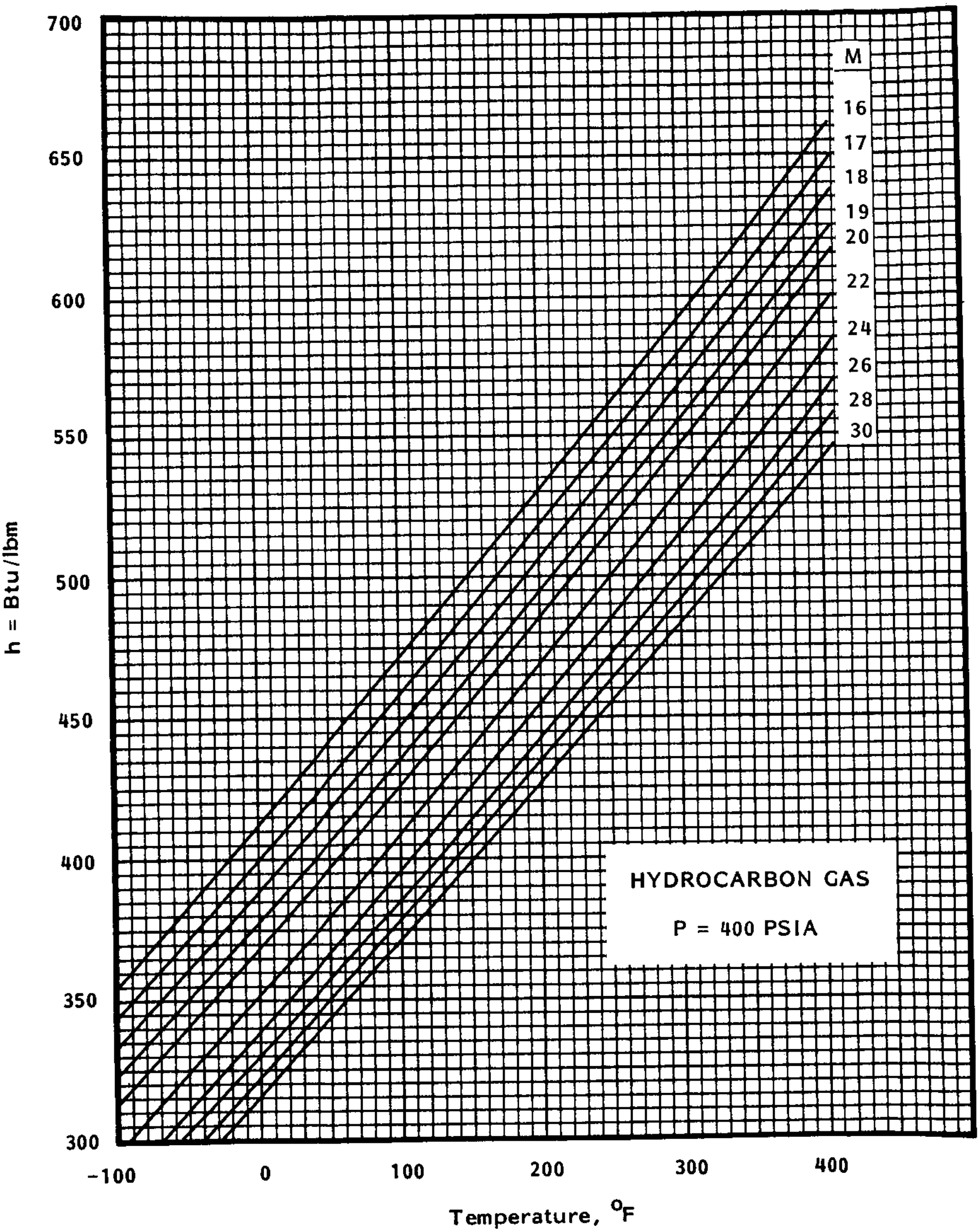

450
400
350
300
250
200
150
100
50
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
30
40
60
80
100
HYDROCARBON LIQUID

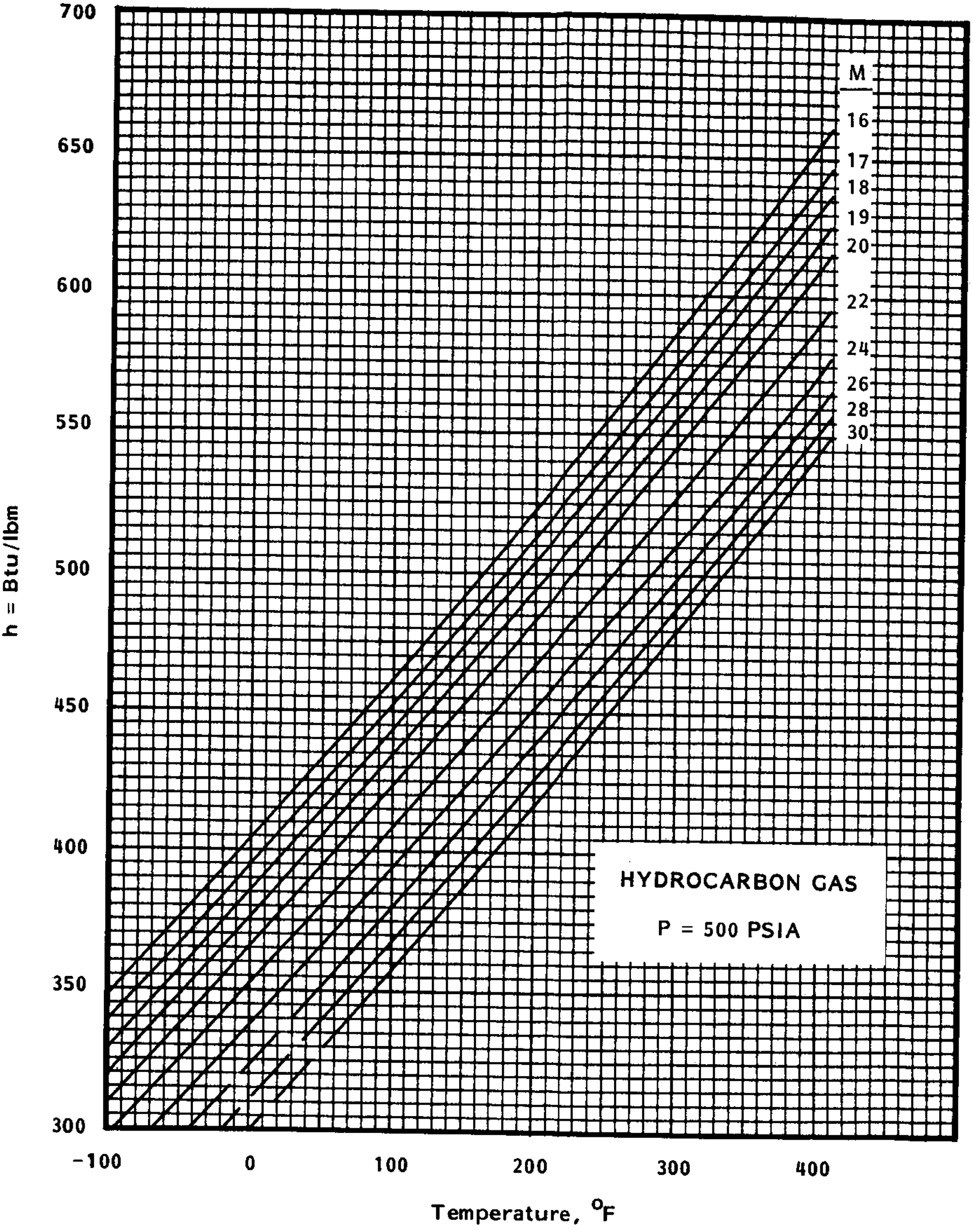

700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 15 PSIA

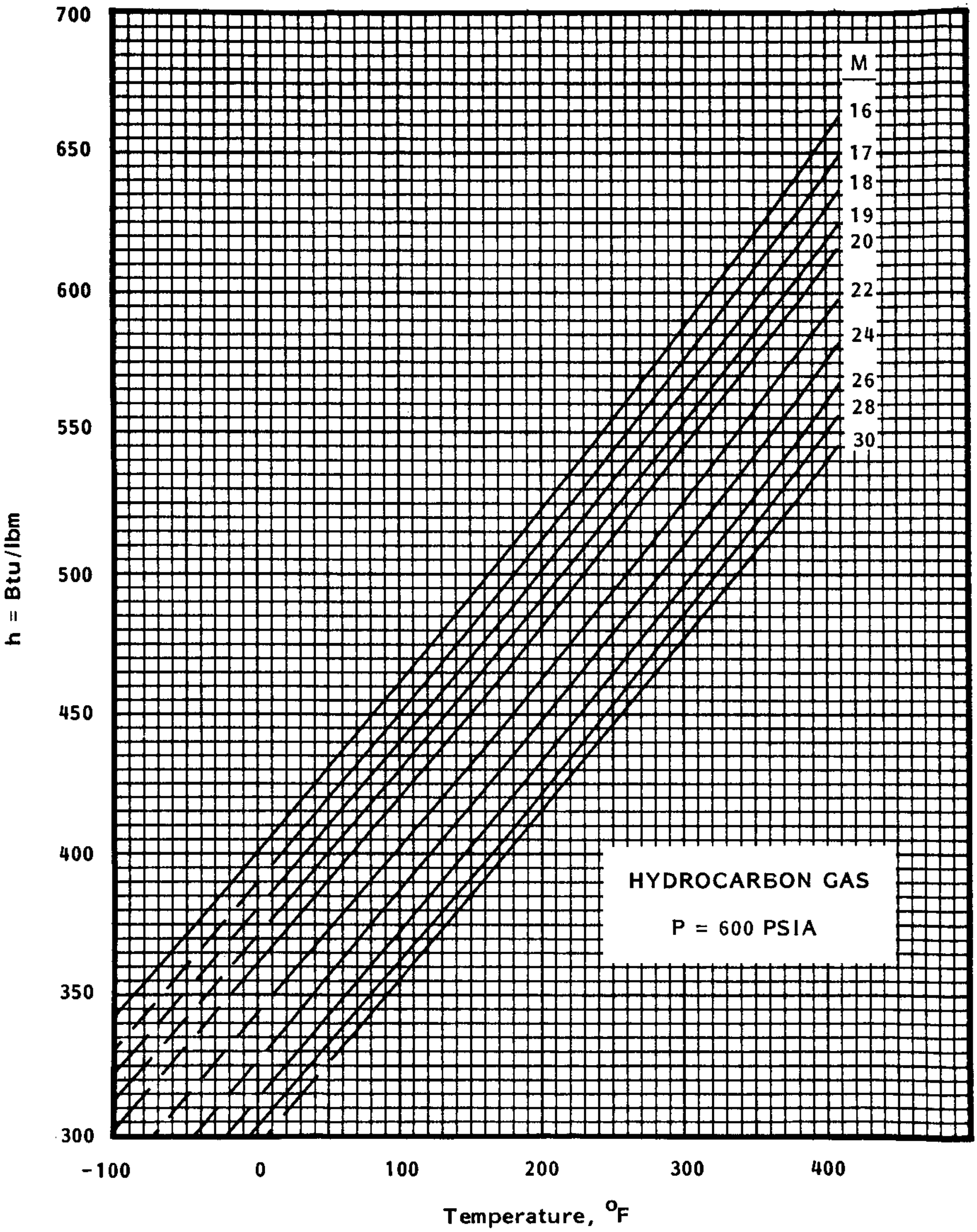

700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 150 PSIA

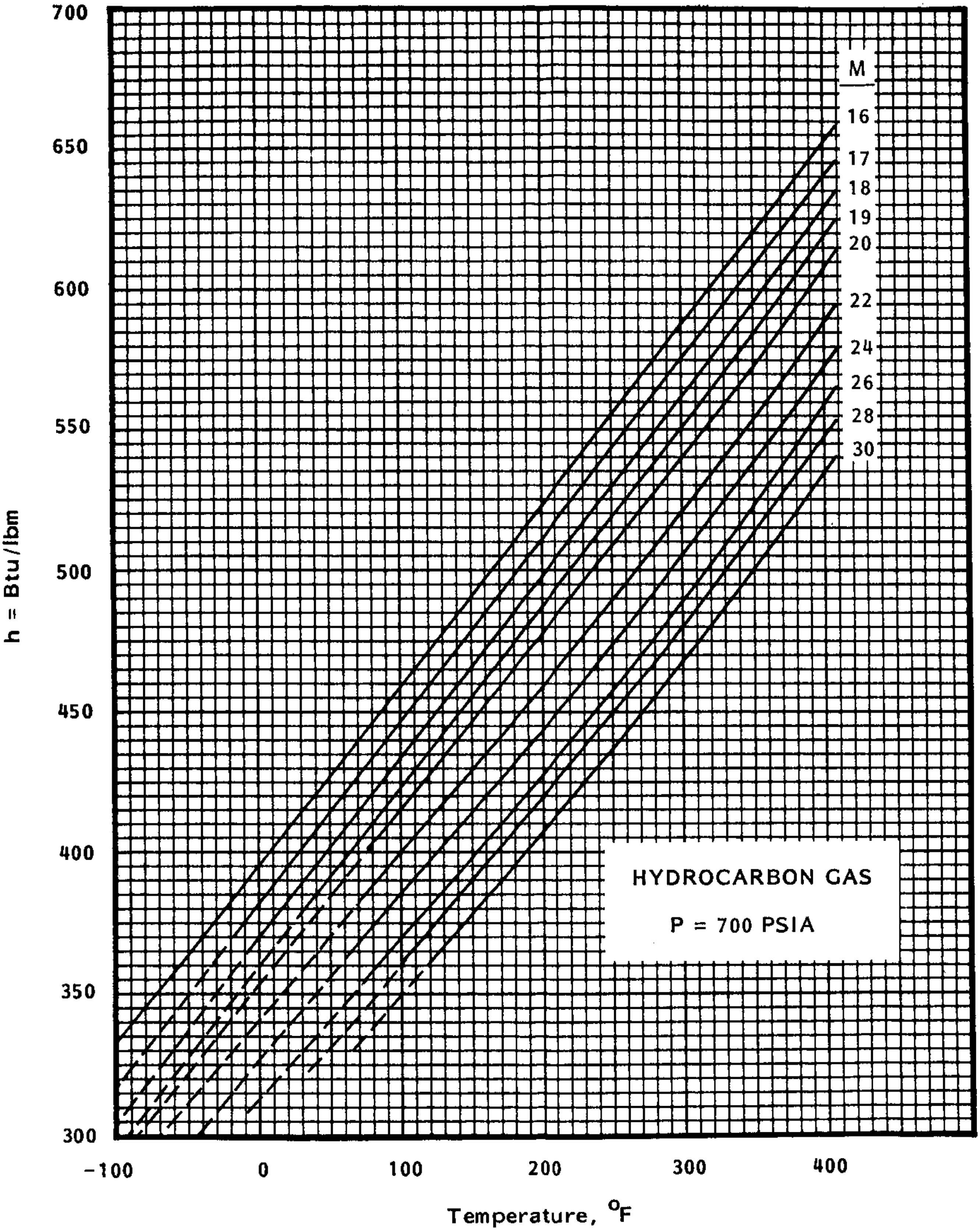

700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 200 PSIA

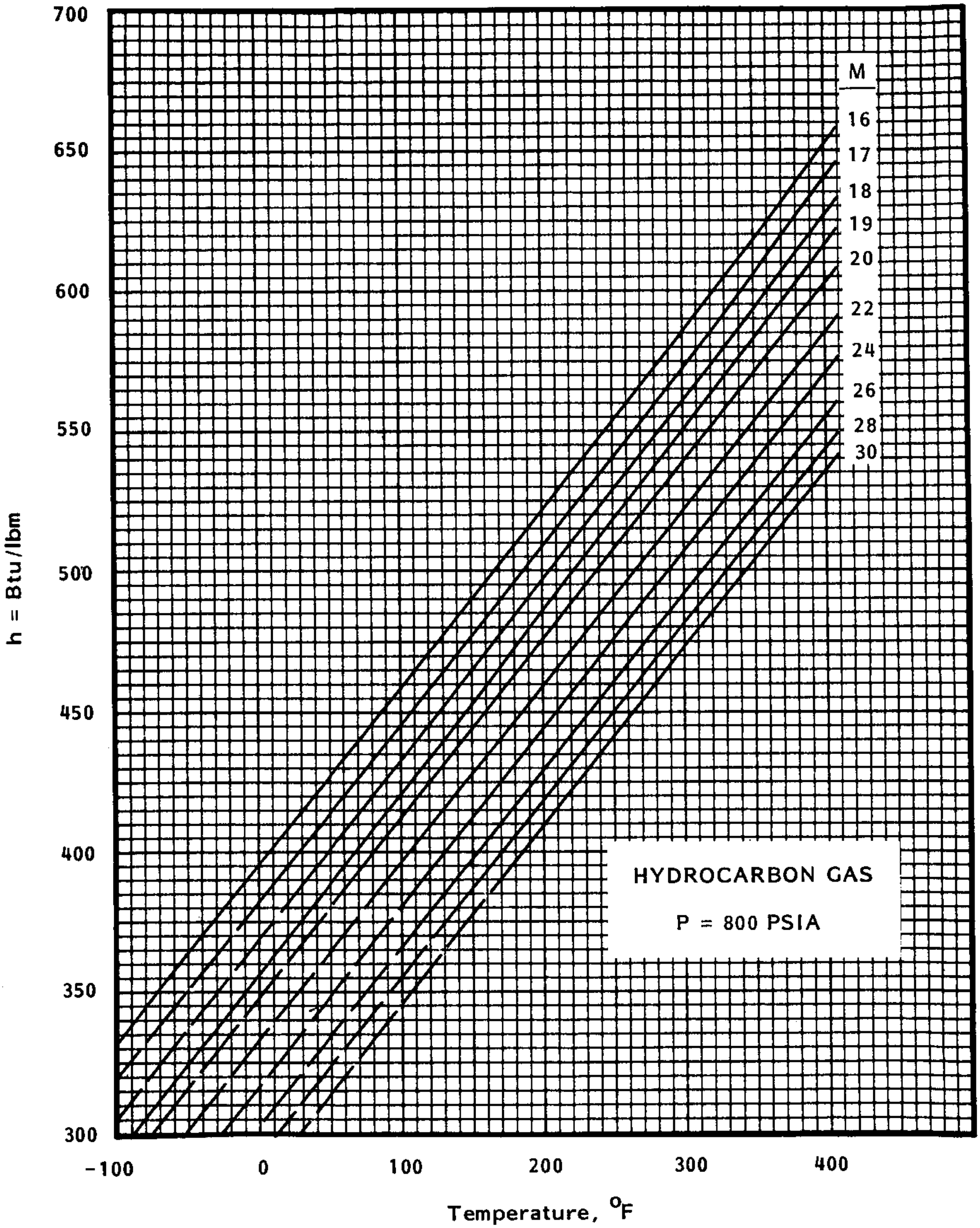

HYDROCARBON GAS
P = 300 PSIA
M
16
17
18
19
20
22
24
26
28
30
h = Btu/lbm
700
650
600
550
500
450
400
350
300
−100
0
100
200
300
400
Temperature, °F

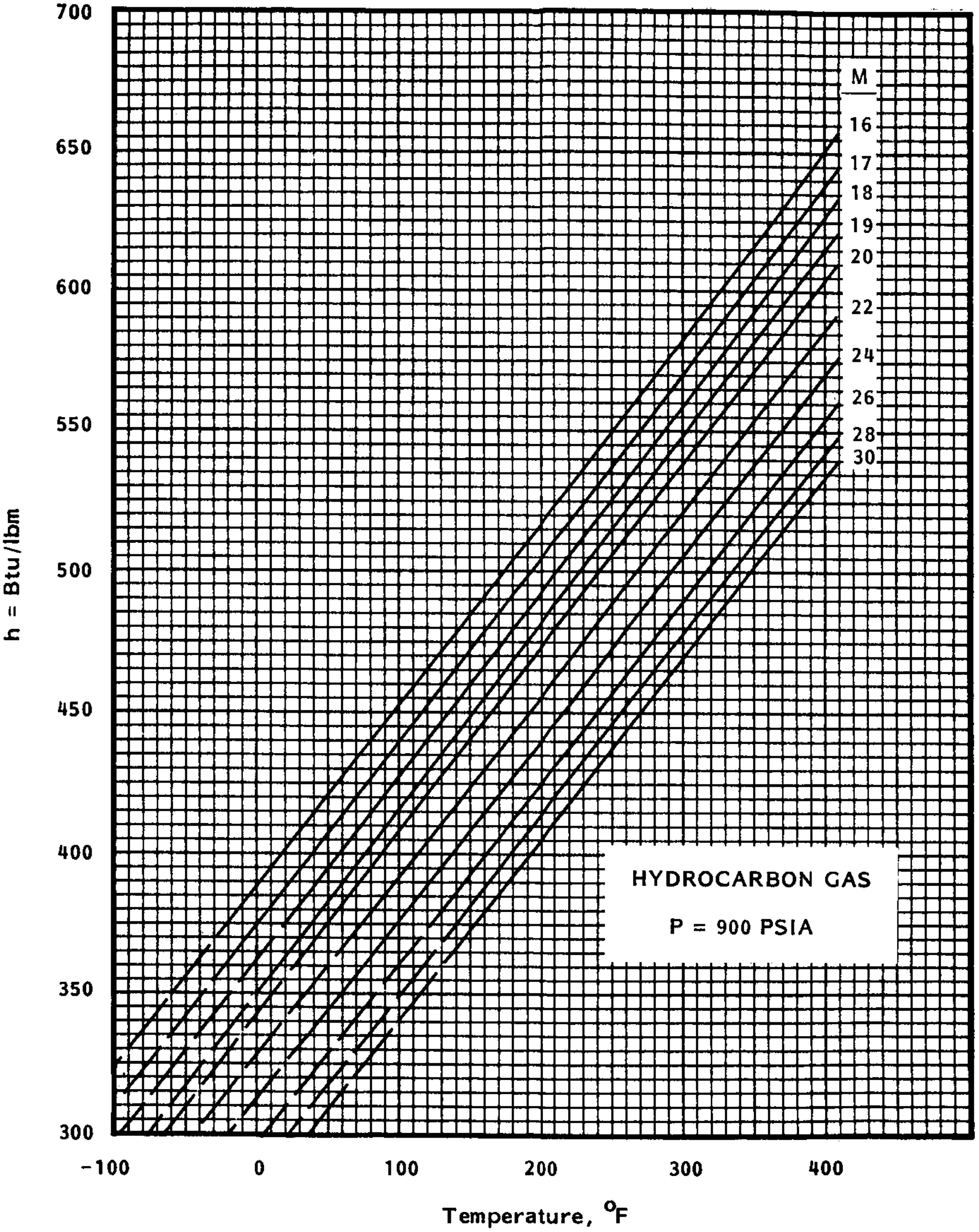

700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 400 PSIA

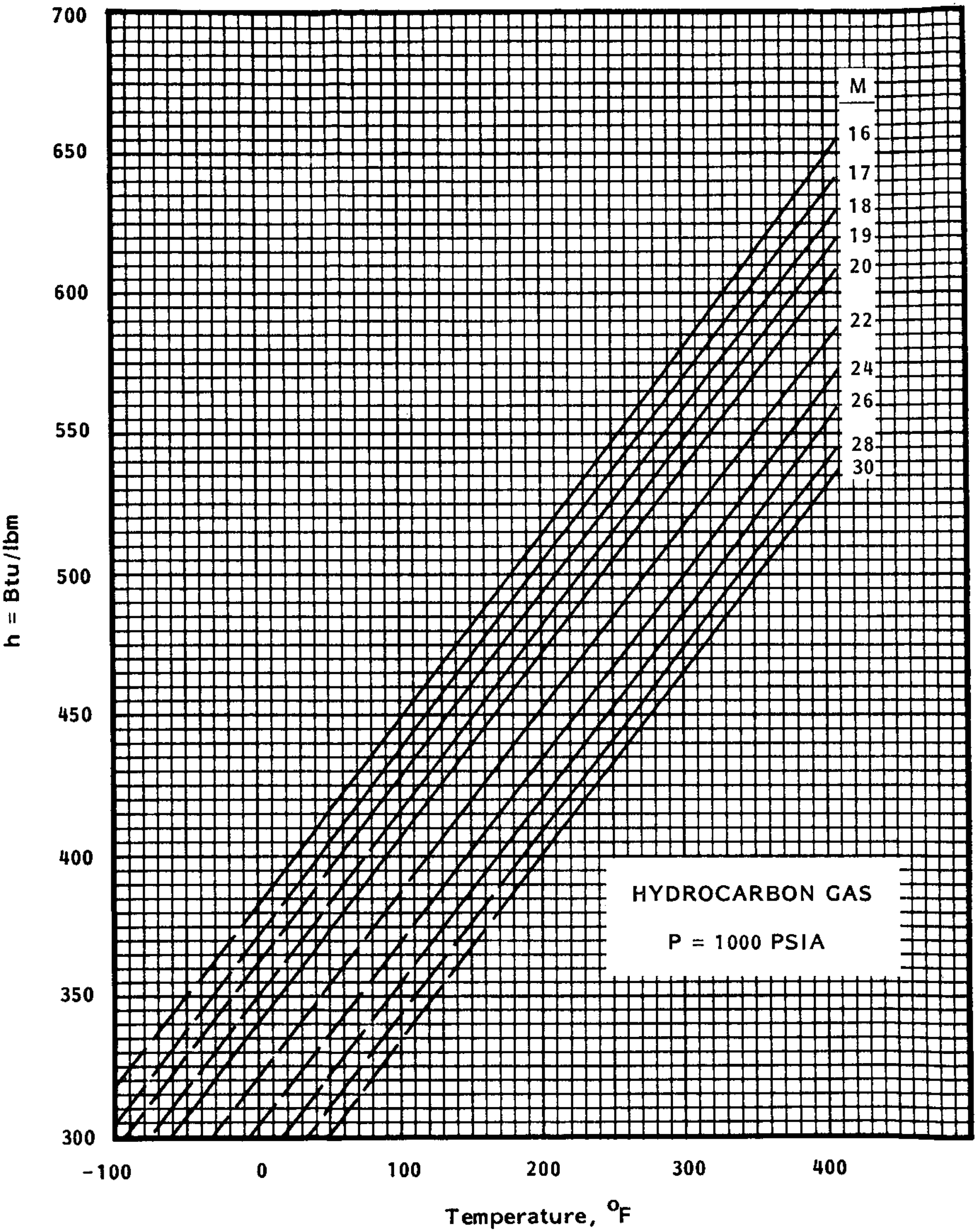

700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 500 PSIA

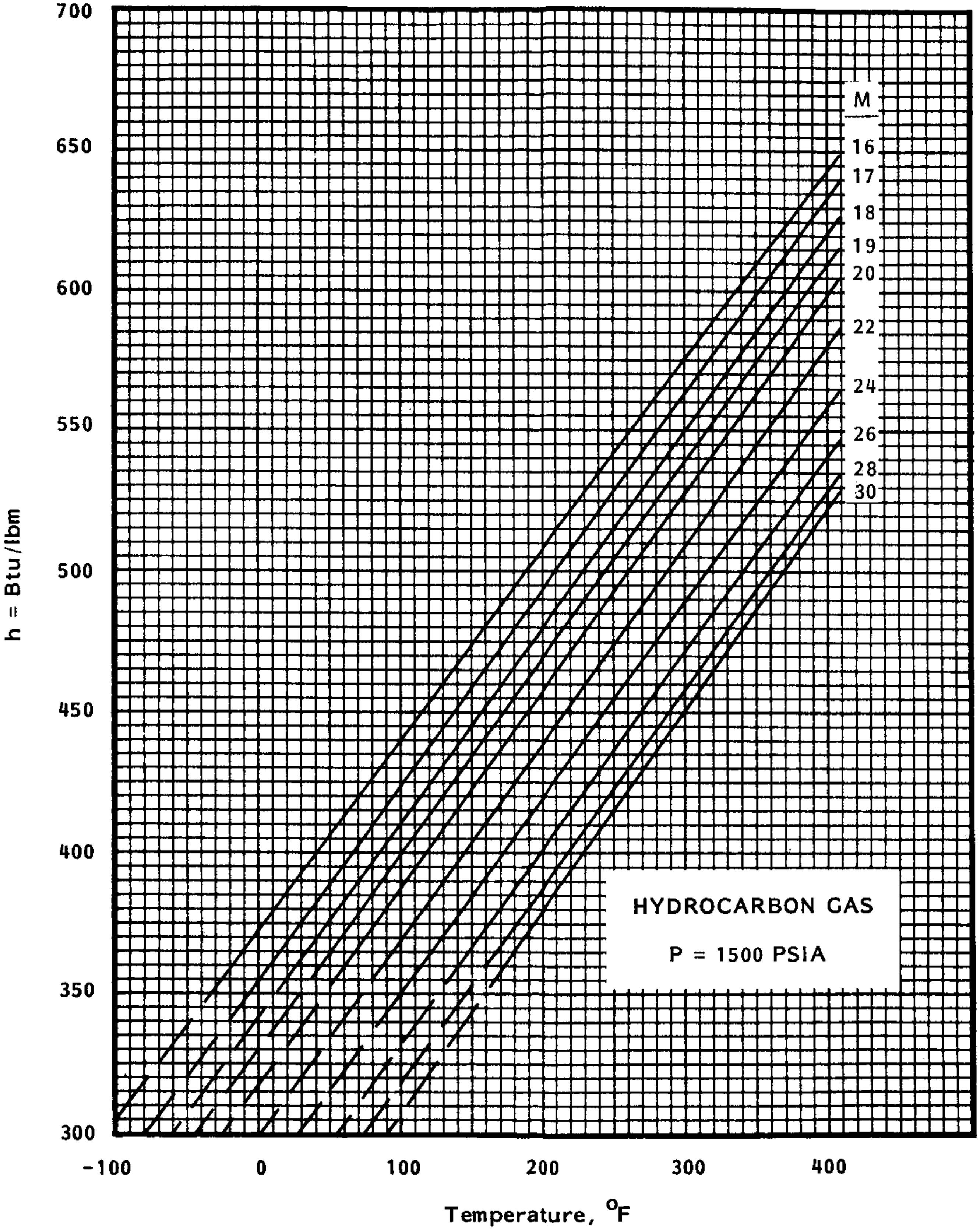

h = Btu/lbm
700
650
600
550
500
450
400
350
300
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 600 PSIA

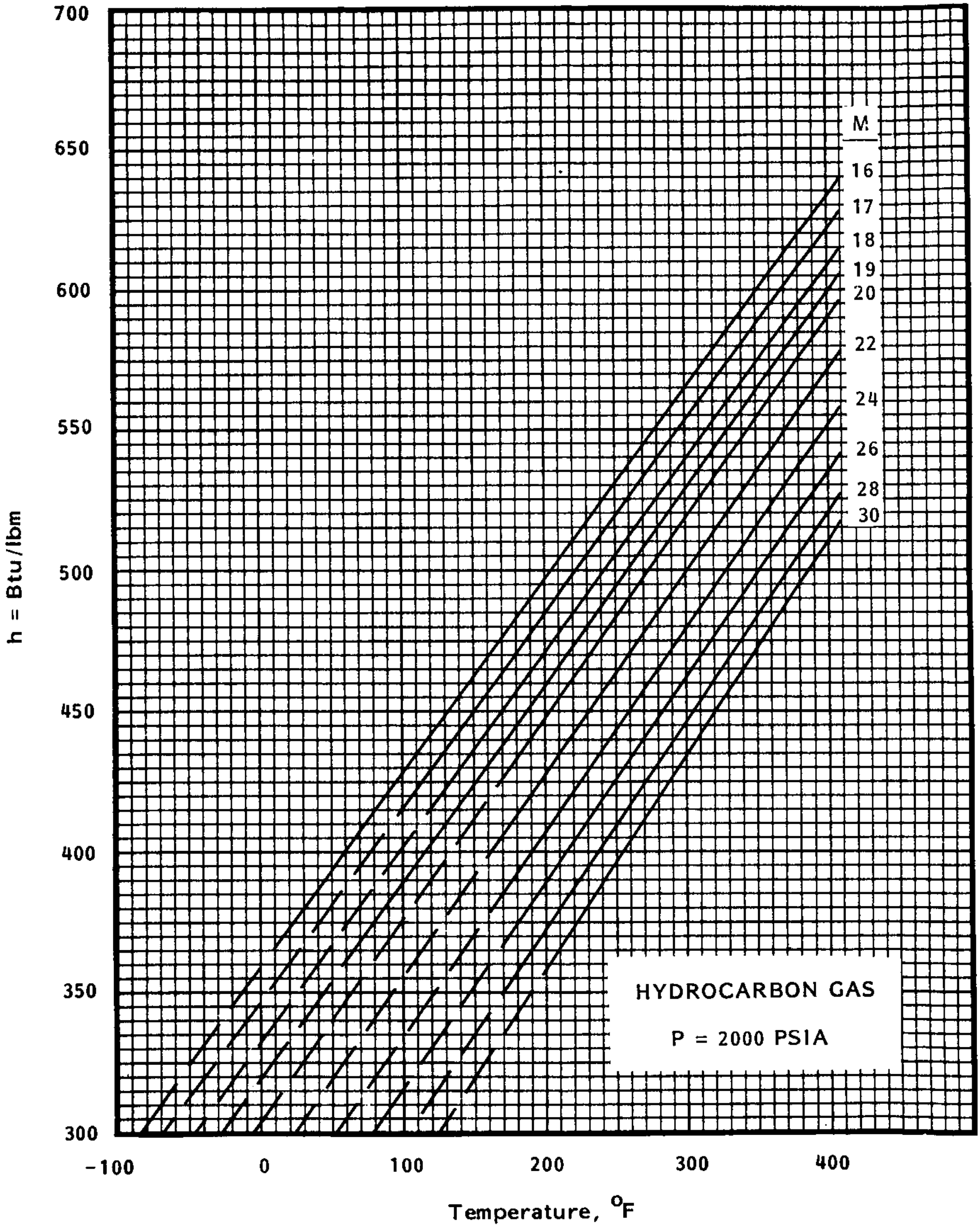

h = Btu/lbm
700
650
600
550
500
450
400
350
300
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 700 PSIA


700
650
600
550
500
450
400
350
300
h = Btu/lbm
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 800 PSIA


h = Btu/lbm
700
650
600
550
500
450
400
350
300
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 900 PSIA


700
650
600
550
500
450
400
350
300
h = Btu/lbm
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 1000 PSIA
-100
0
100
200
300
400
Temperature, °F


h = Btu/lbm
700
650
600
550
500
450
400
350
300
-100
0
100
200
300
400
Temperature, °F
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 1500 PSIA


700
650
600
550
500
450
400
350
300
h = Btu/lbm
M
16
17
18
19
20
22
24
26
28
30
HYDROCARBON GAS
P = 2000 PSIA
-100
0
100
200
300
400
Temperature, °F

9

PROCESS CONTROL AND INSTRUMENTATION

This chapter is an overview of control fundamentals and instrumentation. Specific control system applications are covered in the appropriate chapter dealing with the system design, e.g., compressor control in Chapter 15, fractionator control in Chapter 17, etc. This approach enhances effective communication between the design and instrumentation engineers.

Poor process control is the logical result when this communication is lacking; the process and control responses are not compatible with each other. The process and instrumentation specialists must perform as a team to properly integrate their skills.

The detailed specification of process control is always important, for proper controls are essential if the process is to perform as desired. Even though the vendor of the equipment and control specialists are responsible primarily, the operator must necessarily be conversant with the problem, i.e., understand the principles involved and properly maintain the equipment.

The operator must recognize that the controllers were specified on the basis of an assumed set of design operating conditions. As circumstances change, some change in the control system may be necessary. If the characteristics of the controls plus their related valves are understood, this is not a difficult problem. In many cases readjustment of controls or change of trim size is all that is required.

It has been the author's experience that many problems occur from over- and under-control of the process. The problem with under-control is that an unstable process results. This usually occurs when the installed cost of the facility is critical and/or the vendor minimizes controls for competitive reasons. The only way to guard against this is to visualize the normal process changes that will occur and make a judgment decision jointly with the vendor about the minimum controls needed.

Over-control is a more subtle problem, even though the extra cost is an obvious disadvantage. Too often the assumption is made that the more controls you hang on a piece of equipment, the better the control system. To a point this is correct. Beyond that point, you have controls "fighting" each other. Frequently the superfluous instruments are disconnected and not used in the control system.

Quality and characteristics of the controls can be just as important as their number. For example, a modified inverted bucket steam trap is available as a liquid level controller for small glycol absorbers. It will work and is a bit cheaper than a regular liquid level controller. It is, however, less reliable and more subject to plugging. Which to use is a judgment decision. Does the increased reliability of the unit justify the higher capital cost?

Another important facet is choosing a type of control suitable for the circumstance. A control that might work quite well in a plant using dry instrument air, having good maintenance, and protected from the elements may be totally unsatisfactory in a typical lease production situation. Some compromise is necessary in the latter situation.

The discussions which follow illustrate some of the basic aspects of control. A pneumatic system is used throughout since it has traditionally been used in field processing. Use of electrical controls is quite common today as solid state instruments (including mini-computers) become more prevalent.

The principles discussed can be converted to an electrical system by remembering that pressure and voltage are comparable driving forces.

DEFINITION OF TERMS

In order to properly choose controls one needs to be conversant with both the vendor and field instrument mechanic. A list of commonly used control terms is shown below.

Cascade control – a system in which a process controller controls a second controller rather than a valve directly. The master controller normally resets the set point on a slave controller. Since the slave controller does not have the entire process in its control loop, it can correct rapidly.

Control loop – the system enclosed by the controlling point and the sensing point. In order to achieve proper control, the characteristics and responses of the control system between these two points must be compatible with the process system in the same interval.

Control valve – the device which receives the signal from the pilot and responds by adjusting the manipulated variable.

Controlled variable – the pressure, liquid level, temperature, or flow rate being controlled.

Derivative response – a further addition that provides corrective action based on the time rate of change of the controlled variable.

Diaphragm motor valve – the term applied to a complete valve that uses pressure to open and close it.

Distributed Control System (DCS) – a control concept in which a large control application such as an entire processing facility is divided into multiple smaller subsystems, each which carry out a portion of the application, and the subsystems are allowed to communicate with one another. Redundancy is provided by multiple data transmission highways and the overall integrity of the control system is not jeopardized by the loss of anyone subsystem.

Drift – this item includes reproducibility and the inability to repeat a measurement due to changes in the measured variable.

Dynamic conditions – where the controlled variable is continually changing.

Feedback control – when the output of the pilot is based on the difference between the measured value of the controlled variable and a reference point.

Feedforward control – when the value of controlled variable is maintained by manipulating all the loads (inputs) to the process.

Innervalve assembly (trim) – includes the stem and seat within the valve body that actually controls the flow of fluid.

Manipulated variable – the medium (usually a fluid flow rate) which is manipulated to cause a change in the controlled variable.

Measuring means – the means used to detect any change in the controlled variable.

Normally closed valve – a valve that is held closed by a spring or some similar device and is opened by the action of pilot or the controlled variable or both.

Normally open valve – reverse of the above. One that is closed by the action of the pilot or the controlled variable or both.

Pilot (controller) – a relay that transforms the controlled variable into an equivalent signal to the control valve, such signal controlling the action of that valve.

Pneumatic controls – those which are actuated by air or gas (the most common type used in the field).

Programmable Logic Control (PLC) – a controller which is essentially a minicomputer. It contains the usually proportional, integral and derivative control functions but may also include capabilities for ratio control, deadtime compensation, bias, linearization, valve characterization, etc. It can be programmed by the operator with a CRT and a keyboard.

Proportional control – the system whereby the change in the valve opening is proportional to the degree of change in the controlled variable.

Reproducibility – the ability of an instrument to repeat and measure consistently the value of the static condition over a period of time.

Reset – an addition to a proportional control system to enable the instrument to hold itself at the control point as the process load varies.

Self-operated controller – a valve which is actuated directly by the controlled variable.

Sensitivity – the ability to detect small deviations in the controlled variable.

Servomechanism – a "closed loop" process wherein a "command" received from computer-supplied information is compared with a control feedback which tells what is actually happening. Any difference between these two signals is the error which is fed back into the system for repositioning of the control elements.

Snap-action – a mode of control whereby the valve is either wide open or fully closed.

Static conditions – where all changes in the controlled variable are instantaneous.

Static error – the difference between the absolute value of the controlled variable and the measured value.

Supply gas – that gas necessary in a pneumatic pilot to operate it.

Topworks (motor) – that portion of the motor valve which contains the mechanism to open and close the valve.

CONTROL OBJECTIVES AND CONCEPTS

The objective of a control system is to prevent deviation of a *process variable* from a desired standard. In order to accomplish this objective the control system manipulates an energy and/or material balance. These balances can be summarized in the equation:

$$\text{In} - \text{Out} = \text{Accumulation}$$

The *controlled variable* is usually the accumulation term or surrogate, e.g., level, pressure, temperature, etc. During transient (unsteady state) conditions In = Out and the accumulation variable is changing. The control system maintains the accumulation variable at a desired value by manipulating the appropriate "In" or "Out" variable.

Several factors affect the ultimate control system selected to accomplish the control objective. These include:

1. Nature of the response of the controlled variable (accumulation term) for a given change of the manipulated variable (in or out term).
2. Control stability considerations.
3. Importance of maintaining controlled variable at constant value.

4. System cost and complexity.
5. Penalties (costs) incurred due to poor control.

Feedback Control Systems

The most common type of control system is *closed loop* or *feedback control.* The simple level controller shown in Figure 9.1 is an example of a feedback control system.

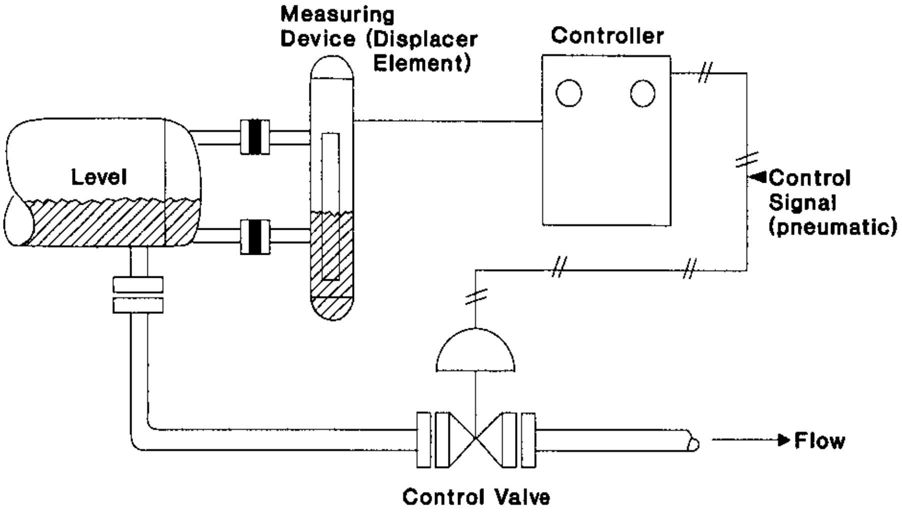


Figure 9.1 Simple Level Control System

In this example, level is the *controlled variable*, and effluent flow rate is the *manipulated variable.* The control signal is generated by comparing the value of the controlled variable (level) to a desired value called *set point.* The difference between the two values is termed the *error* signal. In a feedback control loop the controller output is always a function of the *error* signal. The control signal typically is pneumatic (20-100 kPa, 3-15 psi) or electronic (4-20 ma, 10-50 ma).

The control signal is sent to the *control valve* resulting in a particular valve position or valve opening. Control valves are designed such that a well defined relationship exists between the flow through the valve and the valve position, hence any change in the control signal results in a predictable change in the flow rate of fluid leaving the separator. The changing rate of the fluid flow leaving the vessel causes a corresponding change of the level in the vessel. The dynamics of this change, i.e. (Δlevel/Δflow), are a function of flow rates, vessel size and geometry.

The level control system in Figure 9.1 may also be shown in terms of the flow of information. This is called a *block diagram* and is shown in Figure 9.2.

Process control systems respond to upsets to the process. These are referred to as *load changes.* Load changes may directly affect the controlled variable (level) or the manipulated variable (flow). An example of a load change to the controlled variable is an increased inlet flow to the vessel. An example of a load change to the manipulated variable is downstream pressure from the control valve. Either of these load changes requires a response from the control system. Increased inlet flow rate requires the controller to open the control valve so that Out = In. A decreased downstream pressure requires the controller to close the control valve since flow is a function of valve differential pressure. When the primary feedback loop can not adequately respond to load changes, then more complex control systems may be required. This will be discussed in a later section.

The flow of information in the feedback control loop is not instantaneous. The existence of lags in the process has an important effect on the performance of the control system. A *lag* is defined as the period of

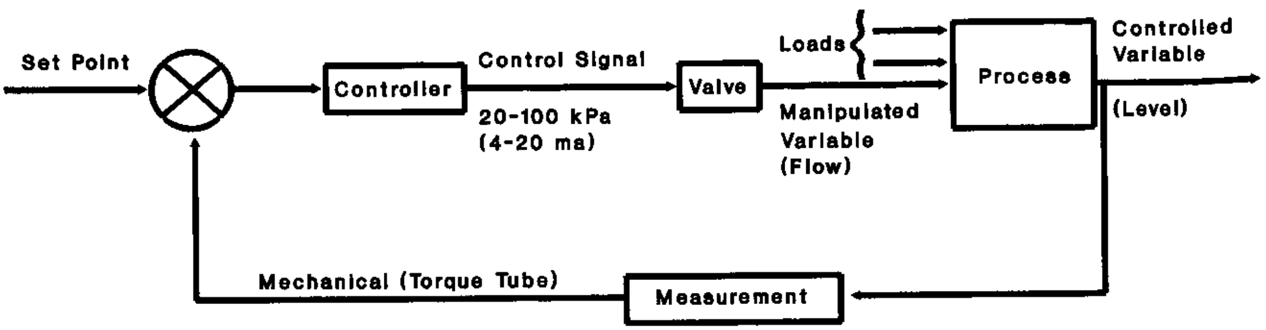


Figure 9.2 Block Diagram for Level Control System

time by which the measured value of the controlled variable follows a given change in the controller output. Lags may be considered in two categories: deadtime and capacitance.

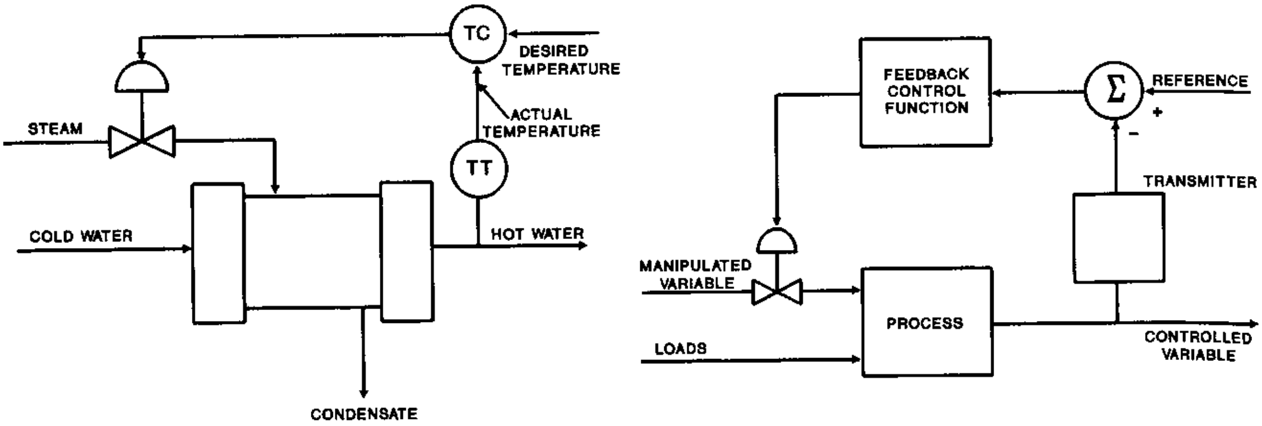
Deadtime is defined as the time delay between change in the control signal and the beginning of its effect on the measured value of the controlled variable. For example, if the flow into pipeline is increased, a period of time elapses before a change in the outlet flow is detected. This elapsed time is a function of the line length and fluid velocity. Because deadtime is caused by the time to move material from one point to another, it is frequently referred to as *transportation lag*.

Deadtime interferes with good control inasmuch as it represents an interval during which the controller has no information about the effect of a load change or control action already taken. When designing a control system every attempt should be made to minimize this delay by properly locating instruments and sampling points, insuring sufficient mixing, and minimizing transmission lags.

Capacitance refers to that portion of the process where energy and/or mass can accumulate. Examples include mass stored in a separator or energy stored in a heat exchanger. The larger the vessel in relation to the flows, the larger the capacitance and hence the more slowly the controlled variable changes for a given change in the manipulated variable. The capacitance of the process tends to attenuate disturbances and hence makes control less difficult.

The capacitance of the process is measured by its time constant. The time constant may be calculated from the differential equation used to model the process, but as an approximation is roughly equal to the process residence time.

These same principles and concepts apply to any control system. Consider the simple heat exchanger in the figure below where water is being heated with steam. Temperature of the water leaving is controlled by the steam flow rate.



a. Process and variables

b. Feedback loop

As with level, this is a reactive system. It doesn't know why the outlet temperature changes; it simply changes steam rate until the set point temperature is obtained. There are three basic characteristics to examine.

1. Control can occur only if the loop is closed. It is "open" when on manual; a sensor or transmitter is faulty or the control valve cannot respond (already full open or closed, or some mechanical fault).
2. Negative feedback is required; the controller must change its output in a direction opposed to the measurement change.
3. Required negative feedback can lead to oscillation within the loop.

Suppose the control valve fails closed. Suppose further that the control loop is set to increase diaphragm pressure with increasing temperature (increase steam flow). This is plus-plus action, the opposite of what is desired. It is called positive feedback. Controller and valve response must be checked to insure negative feedback.

Oscillation can occur in this same heater because of a lag. The outlet temperature lags inlet steam valve movement. The measurement reverses to go toward the set point before the valve can properly respond. Control of oscillation is possible by using different modes of feedback control.

Feedforward and cascade control will be discussed in later sections. In effect these are additions to feedback control which is responsible for the value of the controlled variable. They are used when variations in the loads or manipulated variables to the process vessel change and it is desired to react to these faster.

CONTROL MODES

The manner in which the feedback signal from the transmitter is conveyed to the control valve is very important. The manner in which this valve moves as the controlled variable leaves its set point is critical to good control. Involved is selection of the proper valve and its internals and the controller (pilot). These must be a "matched pair" whose response characteristics are compatible with those of the process.

Controllers

In order to choose controls properly, one must know not only the requirements of the process but also the corresponding characteristics of the controls available. The controller characteristics are determined by the control modes. The controller mode ultimately chosen depends on the process dynamics, control objective and controller cost. The four basic control modes are:

1. On/Off (Snap-Acting)
2. Proportional (Throttling)
3. Integral (Reset)
4. Derivative (Rate)

On/Off

Control is the simplest of all control modes and consequently the least expensive in terms of initial cost and maintenance. A room thermostat is an excellent example of on/off control. In the field, on/off control is referred to as *snap-acting*. Level controls on low pressure separators and temperature controls on indirect fired heaters are often snap-acting. In these cases the dump valve or fuel valve is either open or closed depending on the signal from the pilot. No attempt is made to "throttle" the valve based on the error signal. Snap-acting controllers have the advantage of simplicity, ruggedness and ease of maintenance, and are particularly well suited for remote field locations using wet supply gas.

In many cases the irregular value of the controlled or manipulated variable resulting from a snap-acting control creates serious operating and control problems downstream of the process. In these cases proportional or throttling control is used.

Proportional Control

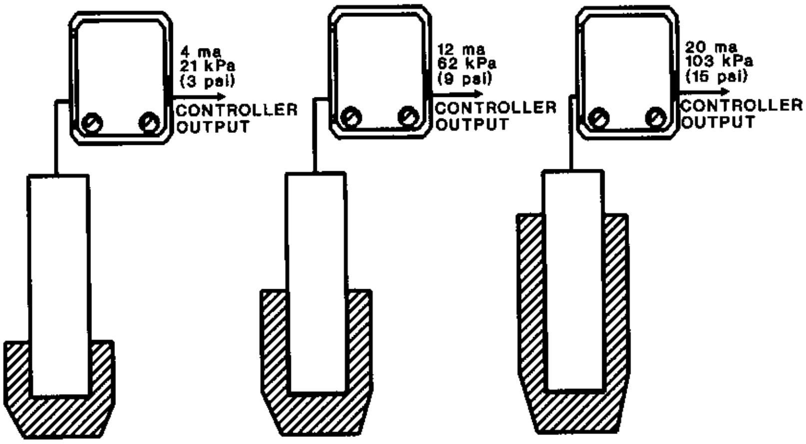
Proportional control is the basic action employed in all controllers not using snap-action. It is defined as control action in which a continuous linear relation exists between the value of the controller output and the error signal. The ratio of change of controller pressure or electric current to a unit change of the error is the *gain*:

$$\text{Output} = K\,(\text{error}) + B \tag{9.1}$$

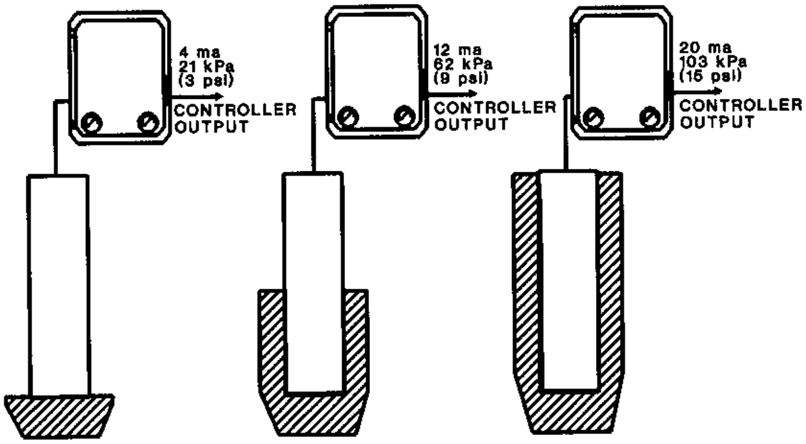
Where: Output = controller output pressure or current signal
K = controller gain, the characteristic sensitivity constant of the controller relating output to input
Error = difference between value of measured variable and set point
B = manual bias

Another way of expressing controller sensitivity is *proportional band.* The proportional band or range, expressed as a percentage, relates the percentage of the full range of the measuring device that the controlled variable has to traverse to fully stroke the valve. It is numerically equal to (100/K).

A good illustration is a typical level-control application shown below utilizing a displacement-type controller, i.e., one which does not float on top of the liquid but depends on the varying buoyancy of the liquid as the level changes.

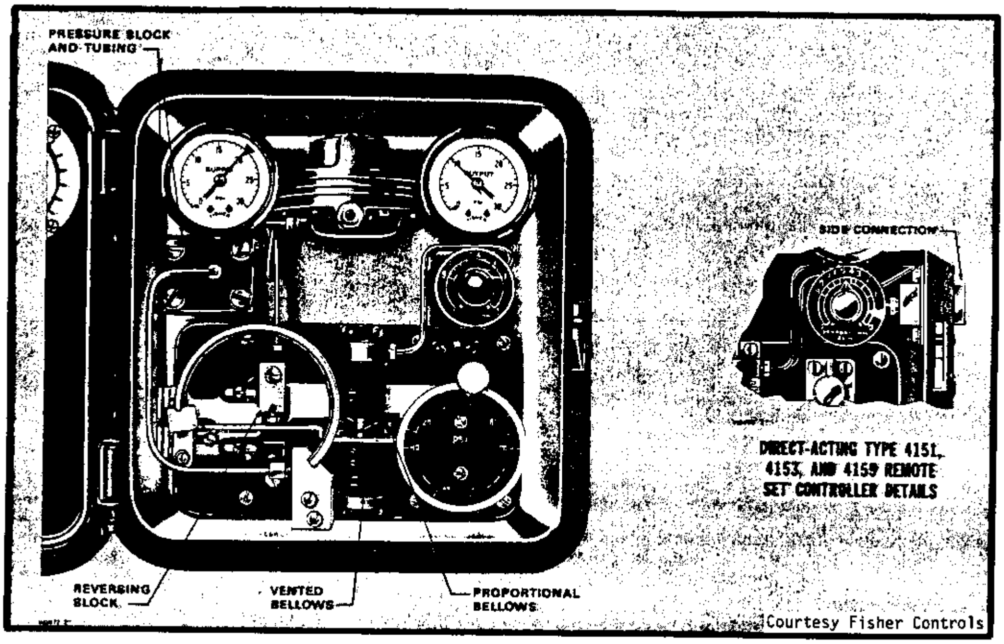


Controller output with 50% proportional band
Controller output changes from 0 to 100% of control range with level change of 50% of float length.



Controller output with 100% proportional band
Pilot output pressure changes from 0 to 100% of control range with level change of 100% of float length.

Note that the narrower the proportional band the more sensitive the control. A proportional band setting of zero is equivalent to snap-acting control: when one adjusts the sensitivity of a pilot the scale is almost always in terms of proportional band. This is clearly indicated in the picture of a common pilot used in plants and production facilities.

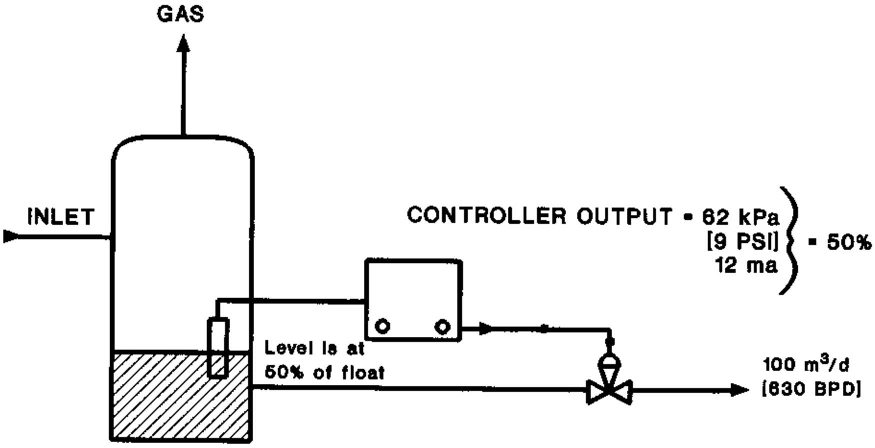


Courtesy Fisher Controls

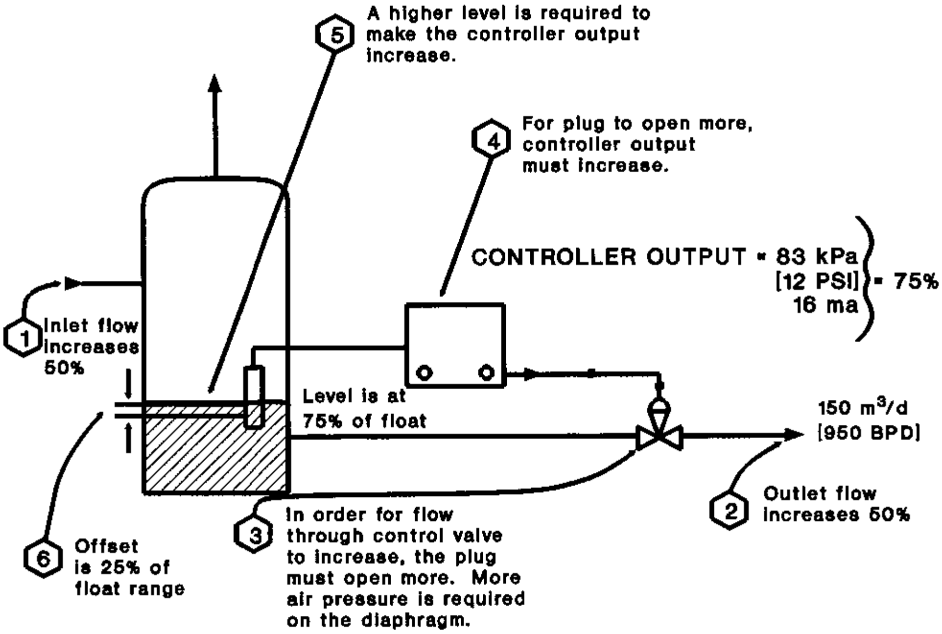
The width of the *proportional band* used depends on the capacitance of the process system, the speed of the corrective action needed, and the limits of control desired.

A narrow proportional band usually is desirable when capacitance is high and fast corrective action is desired. Hence, slow process reaction rate is advantageous to proportional control. Large lag or dead time therefore limits the application of this control by itself. Proper valve sizing for the flow rate involved is a critical factor. The optimum band setting is the least that may be used without "cycling."

In most control systems proportional action is adequate to meet the control objective. However, with proportional action the controller output (hence the valve position) can only change if the measured value of the controlled variable changes. In other words, in the following separator example, the only way the control valve can be moved is if the level in the separator changes. This results in *offset.* Offset is the difference between the measured value of the controlled variable and the set point. The following example illustrates offset.



Level in Separator at Normal Flow



Level in Separator when Liquid Flow Increases 50%

In many systems offset is not a problem. In the preceding separator example, offset would probably not be a serious problem because the objective of the control system is to keep the outlet oil rate equal to the inlet rate. The actual level in this separator is not important as long as it falls between the high and low level alarms.

However, if the vessel is a chiller filled with refrigerant it would be imperative that the refrigerant level be kept above the process tube bundle. In this case offset would be intolerable. Offset can be minimized by increasing controller sensitivity (narrowing the proportional band) but a controller which is too sensitive may be unstable, i.e., swing wildly at the smallest upset.

Reset

Reset or integral action is added to the controller to minimize or eliminate offset. In proportional action, the amount of valve movement is proportional to the deviation of the measured variable. By contrast, the amount of correction applied with reset action depends both on the magnitude and duration of the deviation away from the control point. The prime purpose of reset is, therefore, to prevent offset and keep the controlled variable at the control point even as the process load changes.

The differential equation of a controller with proportional plus reset action is:

$$\text{Output} = K\,[\text{error} + 1/I_t \int (\text{error})\,dt] \qquad (9.2)$$

Where: I_t = integral time, min

In this type of controller the two corrections occur simultaneously. When a change in process load takes place, the controlled variable is returned to the control point with a minimum of cycling. The initial movement of the valve corresponding to a change in the measurement variable is due to proportional action, but the ultimate movement of the valve to its new position is due entirely to reset. As a result, it is often said that reset lags proportional action. The proportional action provides the quick response to correct for the upset, while the reset provides the gradual correction to bring the controlled variable back to the set point.

The reset action is usually expressed in terms of minutes/repeat which represents the time (in minutes) for the reset action to repeat the proportional action.

The reset rate adjustment has to be set by trial and error. When reset rate is set correctly, valve movement occurs at a rate to which the process can respond. If set too fast, cycling will result because the valve moves faster than the process and the measurement, and cycling results. If set too slow, the process will not recover quickly enough.

In a pneumatic system this is accomplished by using a double-bellows system with a suitable fluid between them. The fluid passage contains a needle valve which is the reset control. In an instrument possessing both proportional and reset responses, only proportional is active when this valve is closed. When at the set point, fluid pressure on both sides of the proportional bellows is equal.

Any reset action occurs as soon as the proportional bellows has moved away from the set point position. The reset bellows acts as an integrating device and compensates for this change by the flow of the fluid to or from it, to compensate for any change in the proportional bellows. The rate of such flow is controlled by the needle valve opening.

Derivative

Proportional plus reset control does not provide correction that is rapid enough for certain processes. *Derivative* or *rate* response may therefore be added to "anticipate" a change in process load and transmit a corrective signal to minimize the lag. This action corrects on the basis of the rate of change of the deviation from the set point. This term stems from the fact that the first derivative of the change from the desired standard is incorporated into the control mechanism. The equation for an instrument with proportional plus reset plus derivative response is therefore:

$$\text{Output} = K\,[\text{error} + 1/I_t \int (\text{error})\,dt + D_t\,(dy/dt)] \qquad (9.3)$$

Where: D_t = derivative time, min
(dy/dt) = derivative of controlled variable with time

Derivative action is seldom needed, for proportional plus reset will handle most fairly stable process conditions. Derivative control is proportional to the rate of change of measurement and causes the control valve to reach a given position sooner. The amount of valve position lead is proportional to the rate of change of the measured variable.

Processes having large lags often need this action wherein the valve is moved farther and faster to "get ahead" of the process load change so that proportional or proportional plus reset can take over. If this early rapid movement is compatible with the lag, it will counteract it.

It is said that rate action leads proportional in that it causes the valve to move faster and further than it ordinarily would with proportional action alone.

Derivative action is normally specified only for temperature control systems which are characterized by large process and measurement lags.

The general applications of the various combinations of control action may be summarized as follows:

Proportional – where process time lag is small in comparison to the apparatus capacity such as tank heating, large surge vessels, etc., or where "offset" may be tolerated.

Proportional plus reset – where it is necessary to use a narrow band to prevent "hunting" or overcontrol, and as the frequency and magnitude of the process load changes become greater.

Proportional plus reset plus derivative response – where long time lags and large capacities are combined with large and sudden load changes.

Control Mode vs. Application

Control Mode	Process Reaction Rate	Load Changes		Applications
		Size	Speed	
On-Off: Two-position with differential gap	Slow	Any	Any	Large-capacity temperature and level installations. Storage tanks, hot-water supply tanks, room heating, compressor suction scrubber.
Floating, Single-speed with adjustable neutral zone	Fast	Any	Small	Processes with small dead time. Industrial furnaces and air conditioning.
Proportional	Slow to moderate	Small	Moderate	Pressure, temperature, and level where offset is not objectionable. Kettle reboiler level, drying-oven temperature, pressure-reducing stations.
Proportional-plus-derivative (rate)	Moderate	Small	Any	Where increased stability with minimum offset and lack of reset wind-up is required. Compressor discharge pressure.
Proportional-plus-integral (reset)	Any	Large	Slow to moderate	Most applications, including flow. Not suitable for batch operations unless overpeaking is allowed.
Proportional-plus-integral-plus-derivative	Any	Large	Fast	Batch control; processes with sudden upset; temperature control.

Sometimes the simplest control is the best. In an indirect type heater, for example, the on-off thermostat may be more efficient than a proportional controller. It depends on the burner used and other factors, but on-off control tends to be superior whenever a large mass is being heated or cooled.

PROCESS CONTROL SYSTEM DEVELOPMENT

Early process control systems utilized locally mounted instruments and controllers. Controllers were typically pneumatic and located near the measuring point and the control valve. Operators had to physically go to the controller to change setpoints or take measurements. Figure 9.1 is an example of this type of control system. These systems are still used today in many older onshore production facilities.

As processes became more complex, the use of central control rooms was needed to improve the operation's effectiveness and plant efficiency. Central control rooms also decreased personnel requirements and resulted in better quality control.

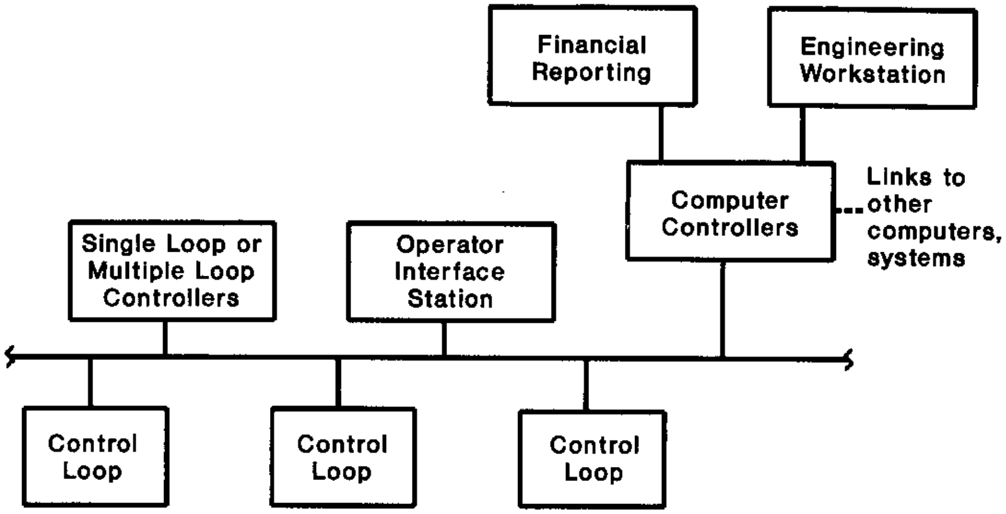
In the early 1960's centrally located mainframe computers were incorporated into the control system to provide some degree of optimization, special algorithms and other central functions not conveniently available from pneumatic systems. Early computer systems were interfaced with the pneumatic controllers. Reliability was usually poor and maintenance requirements were high. Many of these systems provided nothing more than expensive data gathering. Process information could be collected, saved and manipulated but frequently the computer was not "in the loop". The primary reason for this was reliability. Severe problems could occur if the central control computer failed. Back-up computers were typically not used due to their high cost.

In the middle 1970's the first distributed control systems (DCS) were installed. Their introduction was spurred by the need for more reliable control systems and the development of the microprocessor chip.

Distributed control is more of a control philosophy than actual hardware. In a DCS, control functions such as inputs and outputs (I/O), or simulation and optimization routines are "distributed" to the controller. Controllers are microprocessors capable of performing a variety of algorithms on the control signal. These may include the classic PID functions as well as ratio, linearization, cascade, alarm and shutdown. A central computer is no longer needed for such functions, improving reliability, saving money and speeding-up process times.

There are several DCS configurations, but all contain similar elements: remote controllers, programmable logic controllers (PLC's); a communications network often called a data highway; and an operator interface station, usually a keyboard and CRT screen. Most systems also include a main computer for data reporting and interface with engineering workstations or financial reporting.

A schematic of one such system is shown below:



Data transmission is typically digital and redundant data highways are often installed to maximize system reliability. Since many of the complex control functions have been "distributed" to the programmable controller (actually a microprocessor) the entire control system does not suffer if one controller fails. Both single loop controllers (SLC's) and multiple loop controllers (MLC's) are used. An SLC handles only a single control loop while an MLC can handle several control loops and allow information to be transmitted between the loops. MLC's are often used on interactive control systems such as fractionators. SLC's are more reliable since a controller failure only affects one loop.

Another advantage of a DCS is the ability to implement optimization routines into the overall control system. Optimization allows setpoints on various process control loops to be adjusted based on economic considerations.

Algorithms such as simulators, ratio controls, duty calculators can be handled in the distributed controller while the profitability calculations are done in the central computer. In this way as product/energy prices change, the operation of the process equipment also changes to maximize profit. This type of control system has been especially popular for NGL fractionator control.

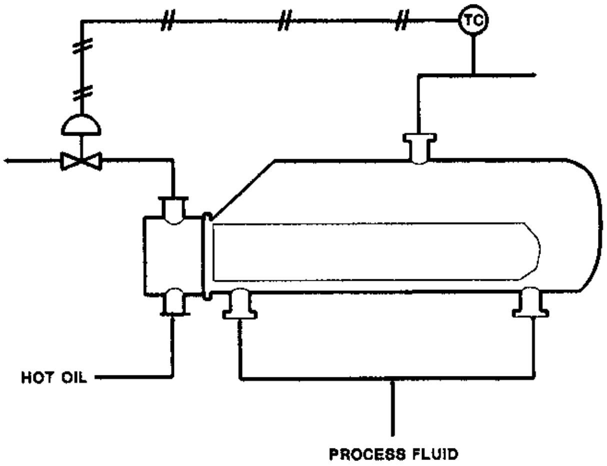
Regardless of the type of control system employed it is imperative that the people involved in the design and operation understand the process. As we have discussed earlier, all control loops manipulate an energy or material balance. Without an understanding of the thermodynamic and rate principles which underlie the process, it is impossible to build an effective control system even with the microprocessor capabilities available today.

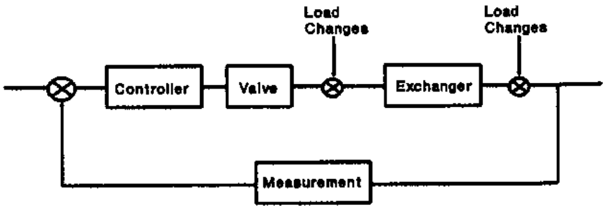
One final caution, while microprocessor based control hardware is more versatile, compact and reliable than older analog systems it requires better trained plant personnel. Companies must be willing to invest the additional monies required to train their process operators and instrumentation staff to effectively use and maintain this equipment. More than one computer based DCS has been derailed by a lack of corporate commitment to technically support the system.

CASCADE AND FEED-FORWARD CONTROL

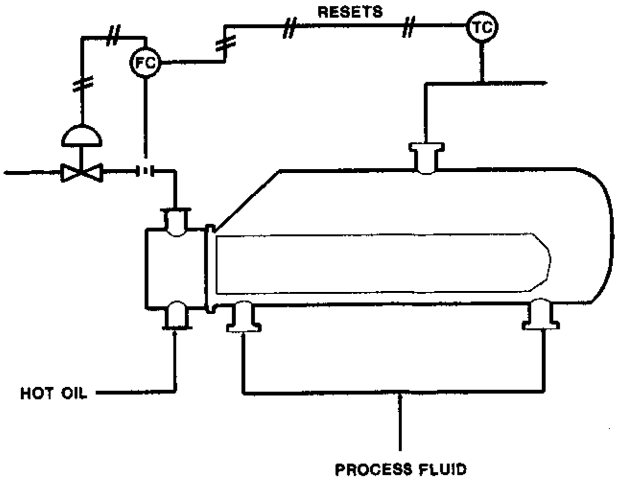
In many cases the simple feedback control system shown in Figure 9.1 is adequate to meet the control objectives. In some instances, however, a more complex system may be required due to the nature or severity of the load changes on the system. This is especially true for sluggish systems such as temperature control.

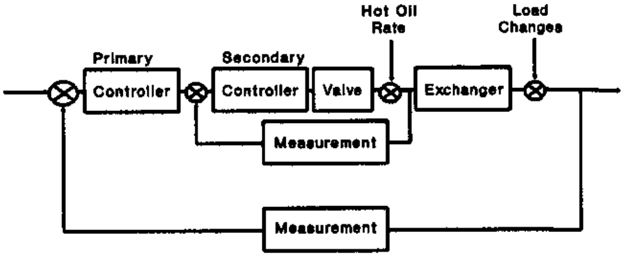
The hot oil system illustrated below is frequently encountered in gas processing facilities. The block diagram for this system is also known.





The temperature controller on the outlet process fluid manipulates the hot oil rate to control the temperature. Load changes may occur on either the manipulated variable (hot oil rate) or the controlled variable (process fluid temperature). When load changes occur on the manipulated variable due to upsets in the hot oil system, *cascade control* is used. An example of cascade control is shown on the next page. In a cascade control system the flow control loop handles the load changes on the hot oil system very quickly without waiting for those upsets to be detected by the process fluid temperature controller. The cascade control system provides superior control to the simple system.

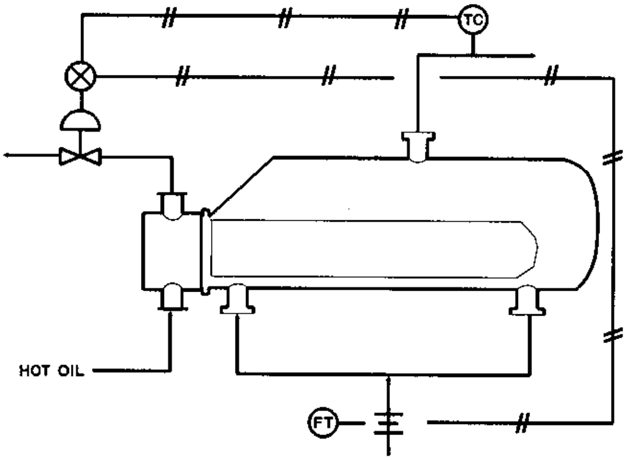


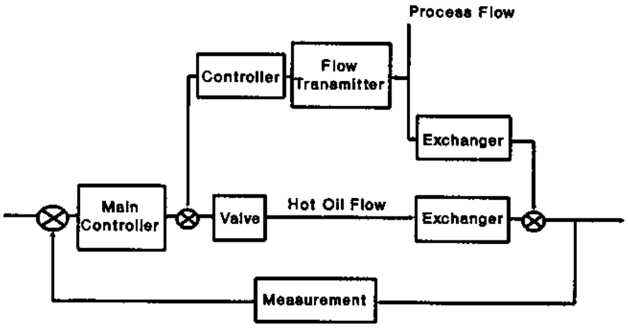


When load changes occur to the manipulated variable and the inside control loop (flow) is much faster than the outside control loop (temperature), cascade control is recommended. Many temperature control systems employ cascade control.

When the load changes occur downstream of the manipulated variable, that is directly to the controlled variable, *feed-forward* control may be required. An example of such a load change in the hot oil/process fluid exchanger is a change in the process fluid flow rate. A feed-forward control system senses the change in flow and sends a signal to the oil valve to either increase or decrease the hot oil rate. Feed temperature could also be measured and the heat duty calculated. The feed forward signal would then be a required heat duty.

Feed-forward control anticipates changes in the controlled variable and adjusts the manipulated variable accordingly. Feed forward control by itself is *open-loop* control – it does not rely on feedback of the measured value of controlled variable. It is almost always used in conjunction with a feed-back control loop often called "feedback trim".. Feed-forward control finds wide application in fractionation and distillation control. An example of feed-forward control is shown below.





CONTROL VALVES

The control valve not only must be matched to the controller but it must be compatible with the pressure, temperature, flow rate and nature of the fluids being handled. Metallurgy, cost, deliverability, parts availability and service are important considerations.

All valves control by varying the area available for flow. Thus any valve that can be moved manually likewise can be moved by controller. Table 9.1 shows some of the valve types commonly used. The choice depends on many factors.

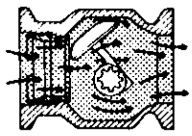
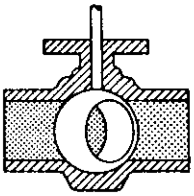
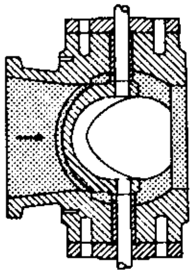
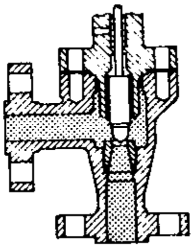
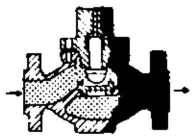
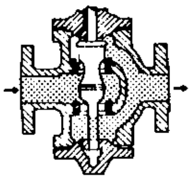
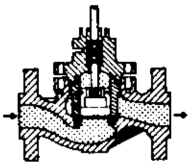
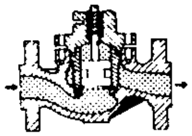
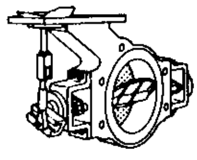
The globe valve following shows the major features of this common type of control valve.

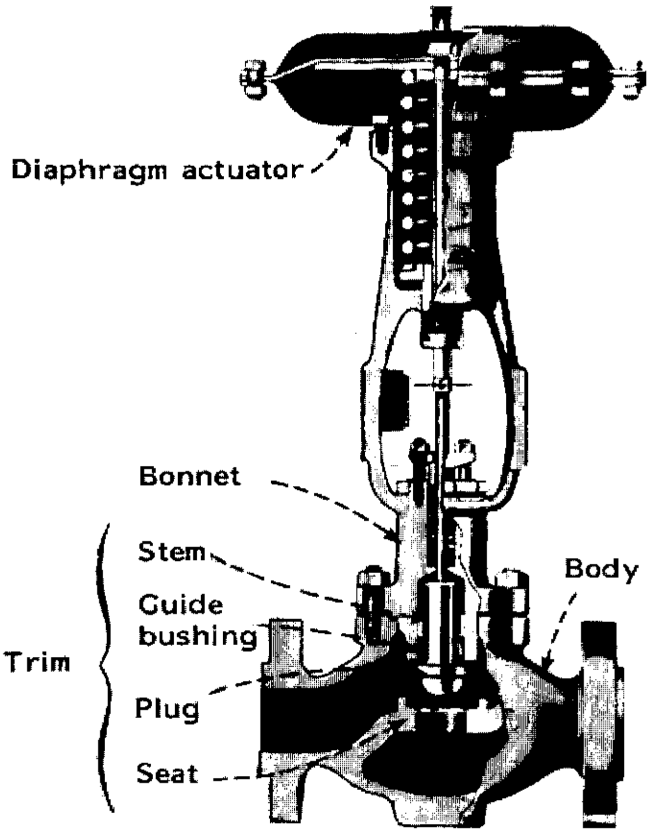
There are three basic parts – the *topworks* (actuator) that drives the *trim* and the *body*. In given models the three are interchangeable to meet both control and process needs.

The plug moves up and down when the unbalance force acting on it is greater or less than the sum of the other two forces involved – the diaphragm force and the spring force. When the upward and downward forces are balanced the plug is stationary.

The spring and plug may be on top of the diaphragm and/or under the seat respectively. A part of this is to properly balance forces and obtain the proper response.

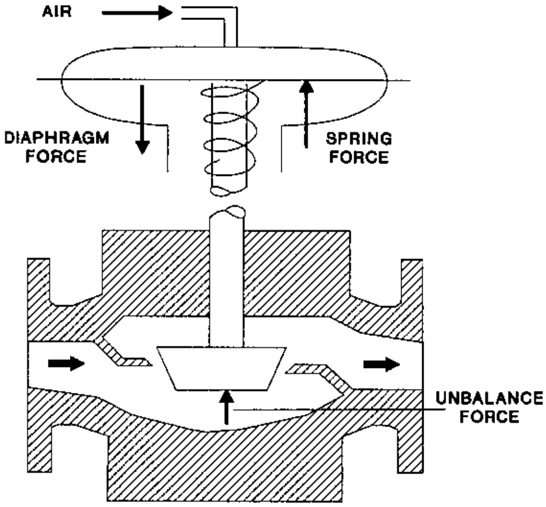
TABLE 9.1
Some Different Types of Control Valves(9.1)

TYPE	
 **Eccentric rotating plug (Camflex)**	Standard sized to 12 in-600 psig; flangeless for inclusion between standard ANSI/flanges, 150, 300 and 600 psig; standard temperature range; –320 to +750°F; equal-percentage and linear characteristics available.
 **Ball valve**	Equal-percentage characteristics; flangeless for inclusion between standard ANSI/flanges.
 **Characterized ball valve**	Standard sizes to 2 in-600 psig ANSI; standard sizes to 12 in-300 psig ANSI; standard sizes to 24 in-150 psig ANSI; all characteristics; flangeless for inclusion between standard ANSI flanges.
 **Angle valve**	Used for special applications; coking hydrocarbons, erosive catalysts.
 **Globe, single port**	Widely available in smaller sizes; standard sizes to 16 in; all characteristics available; standard rating 1,500 psig or 2,500 psig and 450°F; useful in most services.
 **Globe, double port**	Same as single port.
 **Globe, cage, guided, balanced,**	Good availability; all characteristics available; standard sizes to 6 in; standard ratings to 2,500 psig, 450°F; hardened and noise-reducing trim.
 **Globe, cage, guided, unbalanced**	Same as balanced.
 **Butterfly, standard vane**	Widely available in larger sizes and flangeless wafer-style bodies; standard sizes to 36 in-2,500 psig; special sizes to 60 in-2,500 psig, normal characteristic approaches equal percentage; other characteristics available with characterized positioners.



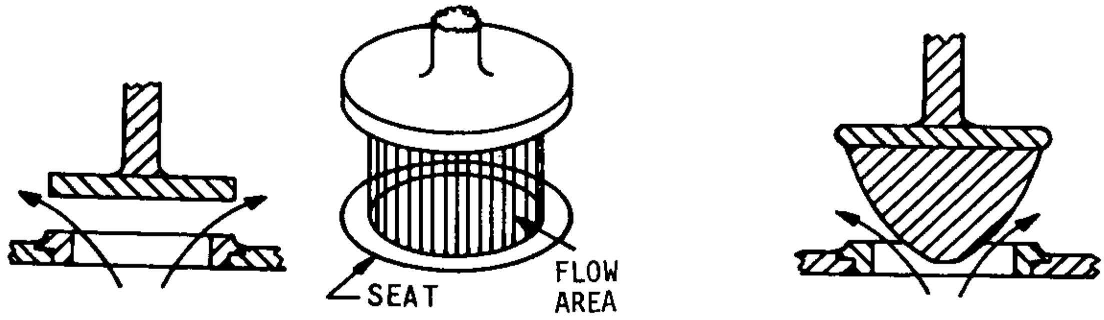
Globe Valve

The valve shown below would be called *normally open* (fail open) because in the event of air (gas) pressure failure, the spring would force it wide open. The valve could be converted to *normally closed* (fail closed) by placing the plug under the seat or by changing the spring in the actuator. The fail position of the valve is an important choice required to allow the system to fail safe.



The diaphragm area, seat opening, plug shape and spring tension must be matched to provide satisfactory control in the service range anticipated. They are not independent factors. One starts with the plug and seat size and shape. Figure 9.3(a) illustrates the effect of stem movement on the variable orifice behavior. With a flat plug the flow area increases rapidly with stem movement. Close control of flow rate is difficult with this quick-opening type of plug. The linear plug, on the other hand, does not increase area so rapidly with stem movement.

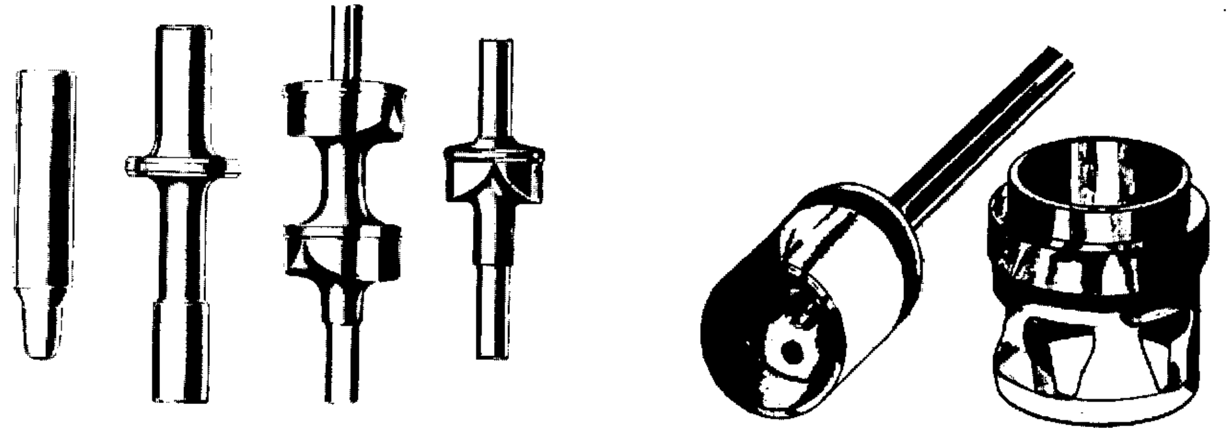
Figure 9.3(b) shows different types of plugs used in different services. Some have an extension of the stem below the seating portion so that the full area of the seat is never available for flow. These are called throttle plugs. Cage guided valves are widely used today. The valve characteristic is determined by the shape of the ports in the cage. Cage guided valves have the advantages of high capacity, flexibility, and employ a balanced plug which minimizes actuator size.



Quick Opening Plug

Linear Plug

(a)



Single & Double Port Valve Plugs

Cage Guided Valve Trim

(b)

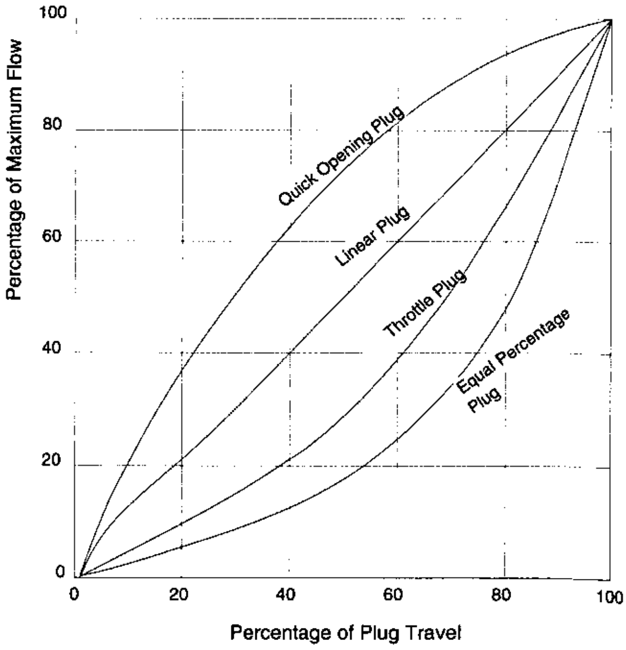
VALVE CHARACTERISTICS

Equal percentage: Denotes a percentage change in flow equal to the percentage change in lift or rotation. Valves having this characteristic are more sensitive to plug- or vane-position change as the valve nears the wide-open position. Since this is normally the operating position, most control valves are equal percentage.

Linear: Denotes equal change of flow for equal plug or vane change at all points.

Quick opening: Denotes a very rapid change in flow, (in some cases as much as 80%) in the first 10% of plug or vane travel.

Other valve characteristics can be produced by varying the shapes and relationships of valve plugs and seats, or by cam or linkage arrangements in valve positioners.



(c)

Figure 9.3 Performance of Valve Trim

Figure 9.3(c) shows the effect of plug type on performance. At axis values 0-0 the valve is fully closed and the flow rate is zero; at 100-100 the valve is fully open and the flow rate is at its maximum. The ΔP across the valve will depend on valve position.

The normal sizing procedure is to select a valve (or trim) size such that over the range of anticipated flows the valve will be between 25-75% open. The valve should not be sized to operate on the seat (less than 10% open) or to operate wide open (greater than 90% open) as this may create stability problems and limit the flexibility of the system.

If ΔP available at the valve is relatively constant over the control range, the desired characteristic is linear. Linear trim is often used in separator dump valves, antisurge valves, fuel gas control valves, etc.

When ΔP varies over the control range, as is the case in pump control, equal percentage trim may be specified because it linearizes the flow vs. travel relationship. Equal percentage trim is also used in systems where the valve has to control at very low rates but also have a very high capacity, e.g., bypass valves.

Quick opening trim is normally used in blowdown, pressure relief and surging systems where high capacity is needed immediately.

Rangeability refers to the valve's ability to "turndown," i.e., handle a wide range of flows. Rangeability is normally expressed as the ratio of valve capacity at 100% travel to valve capacity at 10% travel. Rangeability for linear trim valves is 9-10, for equal percentage valves, 20-40.

In some instances control valves must handle fluids which change phase across the valve. Cavitating and flashing fluids require special consideration in valve sizing and materials selection. In these cases the manufacturer should be consulted.

Flashing fluids are quite common and are generally not a problem. Cavitating flow can cause considerable noise, and cavitation damage can make a valve useless. It must, therefore, receive major consideration in selecting valves for liquid flow.

Cavitation refers to the formation and subsequent collapse of vapor bubbles in the flowing liquid stream. As the liquid flows through the orifice of a control valve, its velocity increases while its static pressure decreases. In many applications, the increase in velocity causes pressure in the control valve to drop below the vapor pressure of the liquid, and vapor bubbles are formed. As the fluid moves downstream within the control valve into a larger flow area, the velocity decreases, with resulting pressure recovery. When the static pressure exceeds the vapor pressure of the liquid, implosions of the vapor bubbles occur, generating extremely high-pressure shock waves that hammer against the valve outlet and piping. Pressures in these collapsing cavities, reported as high as 3450 MPa [500 000 psi] in magnitude, can cause severe and rapid damage to the valve and piping.

Solving cavitation problems begins with either controlling the cavitation process or, ideally, eliminating cavitation altogether.

Erosion-resistant body material, along with hardened valve trim, is a common technique. A sacrificial member, either in the valve itself or in the downstream piping, is another method used to extend valve or piping life. An example is the "flowdown" angle valve, with a hardened liner in the downstream part of the valve body. The success of this method depends on the energy levels involved in pressure drop and flow rate. If both are high, the sacrificial member may soon disappear.

Another approach to preventing cavitation involves sharing the overall pressure drop, either between valves in series or between one valve and a breakdown orifice device or capillary tubing. If the series method is the choice, complicated control problems may occur, and the initial equipment cost may be prohibitive.

Strategic location of the valve along the flow path can give effective and economical cavitation control. For example, locating the valve next to a tank and allowing it to flash into the tank eliminates cavitation problems that could occur if the valve were installed elsewhere in the system.

All these methods have limited applicability or limited design life – perhaps both. In the final analysis, the requirement is usually for a single valve/trim combination that will either control the location of cavitation and prevent damage, or eliminate the potential for cavitation altogether. Most valve manufacturers now design valve trim for cavitating service.(9.3)

Valve Sizing

The sizing of valves requires data from vendors on specific trim. They use a valve *flow coefficient* that depends on the internal dimensions and smoothness of the surfaces. These coefficients are a *capacity index* based on tests made by manufacturers using air or water at an established pressure difference.

In sizing a valve it is normal to calculate flow coefficients, C_{VC}, for normal design flow rate. A valve normally is selected which has a capacity coefficient, C_V, 1.25-2.0 times C_{VC}, depending on the change in flow conditions anticipated.

For dual unit purposes C_V can be regarded as dimensionless but for liquid flow where 60°F water ($\gamma = 1.0$) is flowing at a valve pressure drop of 1 psi, $C_V = q$.

The value of C_V also depends on whether the flow is gas, liquid or two-phase, subcritical, critical, etc. Proper sizing requires a dialogue between customer and vendor. However, Equation 9.4 can be used for routine liquid flow.(9.4, 9.5)

The coefficient C_V is defined by the equation

$$C_V = (q/A)(\gamma/\Delta P)^{0.5} \tag{9.4}$$

Where:		Metric	English
q	= flow rate	m^3/h	U.S. gal/min
ΔP	= pressure drop	kPa	psi
A	= conversion factor	0.0865	1.0
γ	= fluid relative density	–	
C_V	= capacity coefficient	–	

The value of C_V can be supplied by vendors. It is dependent on valve position. The ordinate on the figure in Figure 9.3(c) could be labeled "C_V as a percentage of maximum C_V." Thus the curves shown could be used to estimate a C_V at any valve opening.

The value of C_V varies with the valve but the following values are typical for equal percentage valve trim.

Size, in.	Single Seat	Double Seat
1	8-10	12-14
1 1/4	12-16	16-20
1 1/2	20-22	26-30
2	32-39	44-50
2 1/2	50-57	68-75
3	70-80	105-115
4	120-128	190-210

Example 9.1: It is desired to determine the required C_V for a liquid control valve to handle 100 m^3/h [440 U.S. gal/min] of stabilized condensate at a pressure drop of 690 kPa [100 psi]. γ of liquid = 0.78. From Equation 9.4,

Metric: $C_V = (100/0.0865)(0.78/690)^{0.5} = 38.8$

English: $C_V = (440)(0.78/100)^{0.5} = 38.8$

From one manufacturer's sizing table we find:

Body Size	Travel	C_V @ valve opening, percent of total travel									
		10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
3"	1 1/2"	4.50	5.78	9.96	17.7	26.5	36.6	52.7	71.7	91.5	107
4"	1 1/2"	8.26	13.5	19.9	30.9	44.1	72.5	113	152	182	203

If a 3" valve is used, the valve will operate approximately 62% open, while a 4" valve would operate approximately 46% open. Either would be acceptable although the 3" valve would probably be chosen based on lower cost.

For vapor, the valve sizing equation is written in terms of a C_g and C_1. The term C_g represents the vapor sizing parameter (an equivalent concept to C_V). The C_1 term is called the recovery factor and is a function of the valve geometry.

$$m = A\, C_g \sqrt{\rho_1 P_1}\, \sin\left[\left(\frac{3417}{C_1}\right)\sqrt{\frac{\Delta P}{P_1}}\right]_{Deg} \qquad (9.5)$$

Where:

				Metric	English
m	=	mass flow rate		kg/h	lbm/h
C_g	=	vapor sizing		–	
ρ_1	=	inlet density		kg/m^3	lbm/ft^3
P_1	=	inlet pressure		kPa	psia
C_1	=	recovery factor (C_g/C_v):	butterfly valves	25	
			ball valves	28	
			globe valves	33-35	
ΔP	=	pressure drop		kPa	psi
A	=	conversion constant		0.0457	1.06

The sine term in Equation 9.5 checks for critical flow. If the value of this term (in degrees) is 90° or greater, the valve flow is critical and the quantity in the brackets should be set to 90°.

Actuators

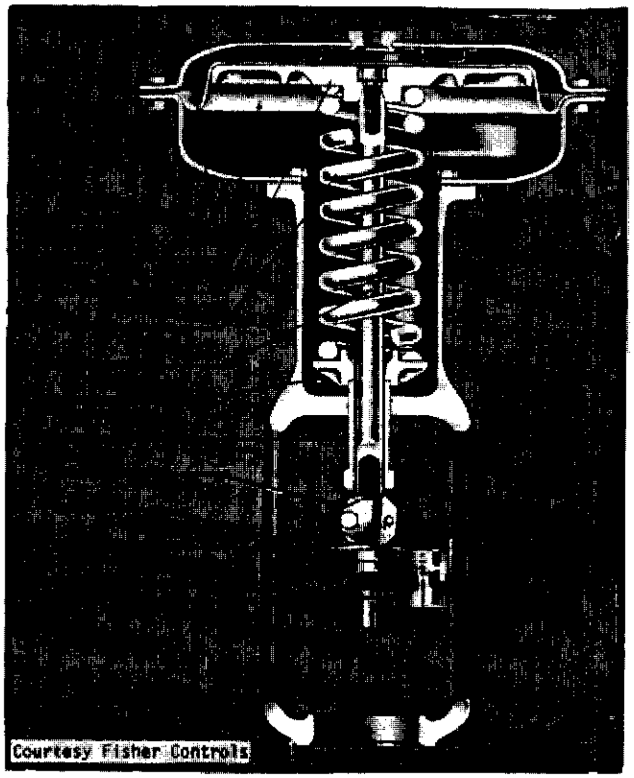
Usually, the final step in valve specification is selecting the valve actuator. An integral part of every automatic control loop, it provides the muscle, or motive force, required to position the valve. And, since stability and operability of the loop hinge on satisfactory actuator performance, the actuator must be able to take control of the many varying static and dynamic forces created by the valve.

For the many available valve styles, there are four basic actuator types suitable for throttling control: 1) Spring-and-diaphragm; 2) Pneumatic piston; 3) Electric motor; 4) Hydraulic or electro-hydraulic.

The pneumatic spring and diaphragm actuator similar to that pictured at right is very common, and offers a simple design, low cost and high reliability.

Diaphragm actuators normally operate with air supply ranges of 21-103 kPa(g) [3-15 psig] or 42-206 kPa(g) [6-30 psig], and hence can receive signals directly from the pilot. The most important advantage of the spring and diaphragm actuator is the ever present capability for failsafe actions. Valves may be selected to fail open or closed on loss of instrument air signal. One of the critical aspects of specifying a control valve is designation of the fail position.

The primary disadvantage of the spring and diaphragm actuator is its limited capability. Much of the diaphragm thrust is used to overcome spring tension and does not result in usable output at the valve. When the thrust requirements exceed 2000 lbf the spring and diaphragm actuator may not be cost effective. In these cases piston or hydraulic actuators are normally specified.

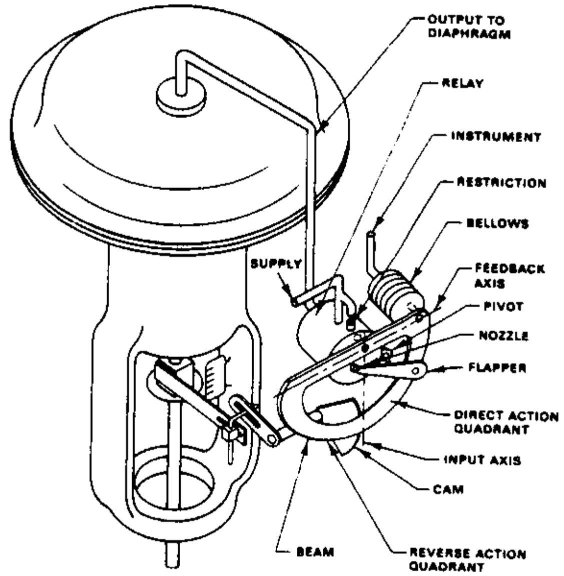

Courtesy Fisher Controls

The table following summarizes the relative advantages/disadvantages of four types of actuators.

SPRING-AND-DIAPHRAGM		PNEUMATIC PISTON	
PRO	CON	PRO	CON
Low cost Simplicity Inherent failsafe action Low supply-pressure requirement Adjustability Maintainability Ability to throttle without positioner Fast stroking speeds possible	Limited torque availability Limited temperature range Inflexibility to changing service conditions	High torque capability Compactness Light weight Adaptable to high ambient temperatures Adaptable to varying valve-torque requirements Fast stroking-speed possible Relatively high actuator stiffness	Failsafe requires accessories Positioner required for throttling Higher cost High supply-pressure requirement
ELECTRIC MOTOR		**ELECTRO-HYDRAULIC/HYDRAULIC**	
PRO	CON	PRO	CON
Compact Suitable for remote applications	High cost/torque ratio Lack of failsafe action Limited throttling ability Slow stroking-speed Lack of adjustability	High torque Very high actuator stiffness Excellent throttling stiffness Fast stroking-speed	High cost Complexity Large size and weight accessories

Valve Positioners

The valve positioner is basically a relay that senses both instrument signal and valve stem position. The positioner applies the necessary signal to the diaphragm to stroke it to correspond with the instrument signal. If the positioner receives a 50% signal, it will apply enough pressure to cause the actuator to stroke 50% of its travel. The positioner will apply the available supply pressure to accomplish this, if necessary.



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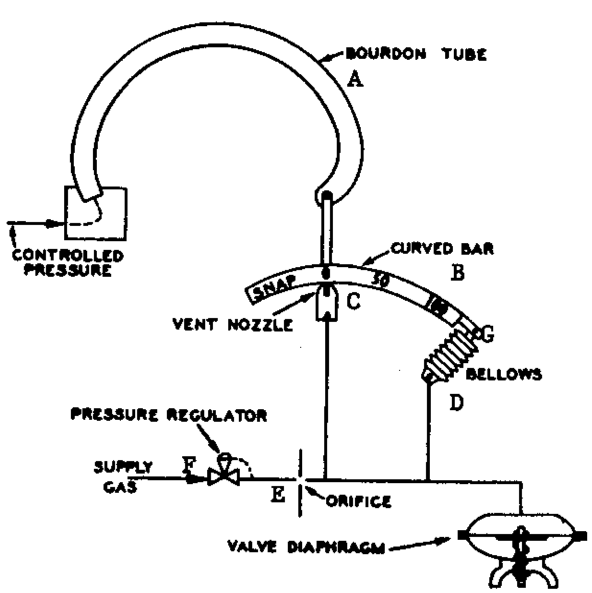
The basic reasons for using a valve positioner or volume booster are:

1. When a split range is required.
2. When a maximum loading pressure greater than the instrument signal is required.
3. When the best possible control is desired for certain systems, i.e., minimum overshoot and the fastest recovery from disturbances.

Valve positioners should not be used as a signal booster. In these cases, one-to-one or two-to-one pneumatic boosters can be purchased at substantially lower cost. When the controller signal is electronic a in valve positioner also serves as a transducer to convert the electronic control signal to a pneumatic signal.

PRINCIPLES OF PNEUMATIC CONTROLLERS

Many pilots still in use in the oil field are pneumatic and use natural gas or air as the actuating fluid. Any natural gas that is free of fluids and at a pressure greater than 150 kPa(g) [21.8 psig] is suitable in this service.



Shown at left is a schematic view of a pressure pilot that will give both proportional and on-off control. As the controlled pressure varies, the bourdon tube (A) will change shape and in turn raise or lower the curved bar (B).

The supply pressure is held constant between 100 and 150 kPa by pressure regulator (F). The vent (C) is so sized that when wide open (curved bar away from it), it will pass more gas than orifice (E). Consequently, the pressure on the valve diaphragm and the valve position depend on the opening of vent (C), which in turn depends on the position of the curved bar, as fixed by the bourdon tube.

If the vent is wide open, the pressure on the diaphragm is zero, while if it is fully closed the diaphragm pressure equals the controlled supply pressure.

It is necessary that both the vent and orifice be very small in order to minimize the amount of gas vented. If the curved bar (B) had a fixed pivot at (G) rather than a bellows, only on-off (snap) action would be possible because of these small openings. From a purely mechanical standpoint, any small movement of the curved bar would in effect make the vent wide open.

The bellows (D) is used to impart a rotating motion to the curved bar around the end of the vent. When the bar begins to rise off the vent, the bellows contract, which tends to keep part of the vent opening covered. As a result, more vertical movement of the bar is necessary to fully open (or close) the valve. All in-between points then represent some degree of proportional control.

The vertical distance that the bourdon tube has to lift the bar to completely stroke the valve increases as the lever arm represented by the distance between (C) and (G) decreases. This then allows for adjustment of the proportional band.

If the vent is close to the left end of the curved bar, the movement of the bellows imparts little rotation to the bar and on-off action results.

Most pressure pilots have the bar marked to show various percentages of proportional control. The percentage shown indicates that the controlled variable must vary through that percentage of the instrument's range to fully open or close the valve.

Example 9.2: If a pressure pilot has a bourdon tube with a range of 0-2 MPa and it is set on 50 percent proportional band, how much must the pressure vary to make the valve be fully stroked?

Solution: It must vary $(0.50)(2 - 0) = 1$ MPa to fully stroke the valve.

It therefore follows that the closer the vent is to the bellows, the higher the percentage of proportional control obtained.

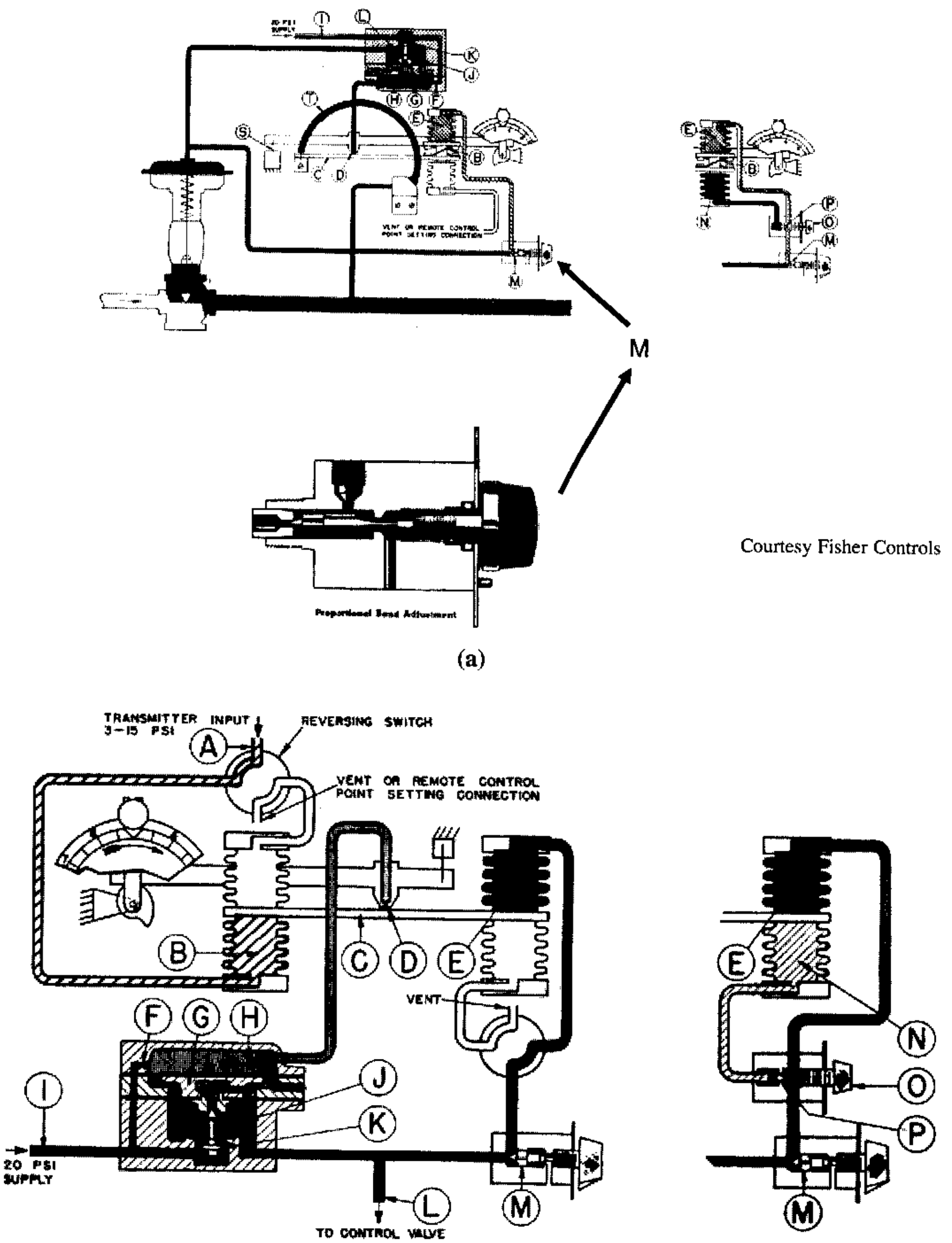
Reset could be added to the controller shown above by adding a second bellows that worked in opposition to the one shown. The two bellows would be connected by a needle valve to control the bleed rate back-and-forth between the bellows. Adjustment of this needle would constitute reset adjustment.

Most pressure pilots are slightly more complex than the schematic presentation shown, in order to improve the mechanical action and the sensitivity. However, the basic mechanism and principle of operation shown are common to all pneumatic pilots.

Figure 9.4 shows a comparable approach in Fisher Controllers. Both pilots are exactly the same except for the manner of input. In the top setup a bourdon tube is used to transmit line pressure to the instrument. In the lower figure a transmitter uses a low pressure signal to activate the controller through bellows. In all other aspects the systems are identical. At the right is shown the bellows/valve addition to add reset to the basic proportional controller.

An increase in bourdon tube or transmitter pressure moves plate (C) towards nozzle (D). This will build up the pressure in chamber (H) of the relay because a constant air or gas supply pressure connected at (I) is maintained on the orifice (F). This buildup of pressure in chamber (H) pushes the relay diaphragm assembly (G) upward opening relay supply valve (K). This allows supply pressure to come into chamber (J) until it pushes the relay diaphragm assembly (G) back to its original position and valve (K) is closed again. A decrease in pressure in chamber (H) will cause the diaphragm assembly to move downward and to open exhaust valve (L), allowing pressure in chamber (J) to bleed out through side of relay.

The ratio of the two diaphragm areas in the relay is 3 to 1. Hence a 5 psi change on large area diaphragm (G) results in a 15 psi change in pressure to the diaphragm control valve. The relay is of the



Courtesy Fisher Controls

(a)

(b)

Figure 9.4 Schematic View of Two Fisher Pilots

intermittent bleed type and wastes air or gas to atmosphere only when pressure to the diaphragm control valve is being reduced.

It is shown schematically that an increase in output pressure from the relay chamber (J) goes to the diaphragm of the control valve and at the same time, goes to a 3-way valve assembly (M). This 3-way valve is the proportional band adjustment which can be manually changed to get the amount of feedback desired to the proportional bellows assembly (E). When the 3-way valve (M) is fully open, all of the diaphragm pressure is transmitted to bellows chamber (E). This causes the beam (C) to move away from nozzle (D) thus stopping the buildup in relay chamber (H), resulting in full proportional band based on the rating of the bourdon tube. Intermediate positions of the 3-way valve (M) will give lower proportional band responses.

If the controlled pressure decreases, the beam moves away from the nozzle (D), causing relay exhaust valve (L) to open. The diaphragm pressure decreases an amount governed by the proportional band setting, causing the main valve to open more and restore pressure to desired setting.

With reset added the output from the proportional 3-way valve (M) goes to bellows (E) and also through another 3-way valve (O) and into the bellows (N). An increase in pressure in proportional bellows (E) causes the flapper to move away from the nozzle. However, with reset action incorporated, an increase in pressure occurs in bellows (N) tending to nullify the action of the proportional bellows and bring the flapper closer to the nozzle. The reset action will cause a definite increase in controller output pressure per unit time interval until the control point is restored to the desired value. On a pressure reducing application, when the controlled pressure decreases, the reverse of the above takes place and reset action decreases the controller output a certain amount each time interval.

Regardless of the manufacturer the basic principles represented by these examples apply to all controllers.

LEVEL MEASUREMENT AND CONTROL

Measurement and control of liquid level is an integral part of both plant and field processing. In view of the widespread applications and variety of conditions encountered, it is not surprising that a number of mechanisms are used. Those prevalent in oil and gas production and processing are:

1. Mechanical or pilot operated valve actuated by a float.
2. Diaphragm control valve actuated by a displacement "float."
3. Diaphragm control valve actuated by a DP cell.

The choice of control mechanism depends on the pressure, flow rate and service needs. In a horizontal separator it may be necessary to hold a very constant level, more constant than in a vertical separator. In some vessels you want the level to rise and fall to dampen out fluid surges.

The simplest level control system uses a *float* that rises and falls with the level; it truly floats on the liquid. The simplest of these is a float operated mechanical oil valve where float movement trips a snap-acting valve through a linkage. Up to pressures of about 800 kPa [115 psia] these are a reliable, rugged piece of equipment. At higher pressures the force required to stroke the valve exceeds the float capability.

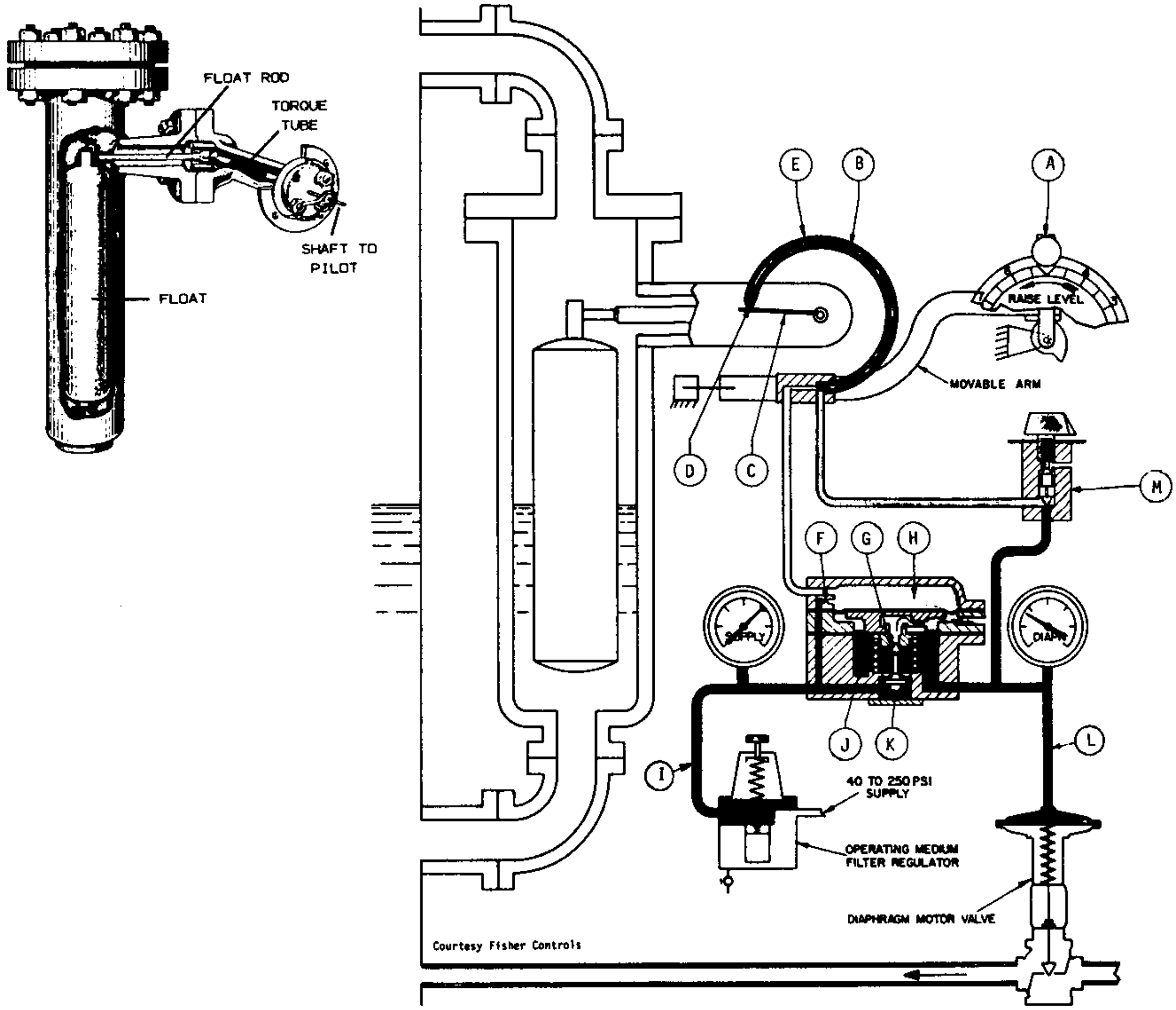
At pressures to about 5.0 MPa [725 psia] this float normally will be combined with a pilot operated valve. This has been a standard unit on vertical separators in oil separation applications. It is a simple system but has limited rangeability and may not work too well in sensitive level control applications.

The *displacement* "float" is the most commonly used level control in production operations. The displacement type float is not a true float since total movement usually is less than one-fourth inch. It operates on Archimedes Principle which states that a body will be buoyed by a force equal to the weight of liquid displaced.

This displacement element usually is about 7.6 cm [3 in.] diameter. For ordinary level service it is 36-38 cm [14-15 in.] long and weighs about 2 kg [4.5 lbm].

In the "standard" control the float is attached to a torque tube by means of a float rod. The torsional force is the weight of the float times the length of the float rod. A small flapper is attached to the end of the torque tube. It moves with change in torsional force to effect the bleed rate from the pneumatic pilot and thus the pressure on the valve actuator diaphragm.

Figure 9.5 is one arrangement using the same basic mechanism shown in Figure 9.4. The flapper (C) on the end of the torsion tube moves in relation to vent (D) which is in the end of bourdon tube (E). When the movement of the bourdon tube and flapper combine to shut off vent (D) the full supply pressure is



Courtesy Fisher Controls

Figure 9.5 Schematic View of Fisher Dislacement Type Level Controller

available for line (L) and the inlet to proportional valve (M); the pressure at (M) and (L) is essentially zero when vent (D) is fully open because it is larger and can pass more gas than inlet vent (F).

Valve (M) is the same one shown in Figure 9.4. It can be adjusted so part of the pressure at (L) can be bled into bourdon tube (E). Any change in pressure causes this tube to expand or contract (away from or toward the flapper). The combined action of the valve and bourdon tube is equivalent to that of the valve and proportional bellows in Figure 9.4.

The system in Figure 9.5 may be used as a *transmitter* to another controller. In this case line (L) is the input to this second instrument which, in turn, is connected to a valve. It could be the source of pressure point (A) in the lower instrument of Figure 9.4. If used as a transmitter the level control in Figure 9.5. would be set at 100% proportional control using valve (M).

The combination of float length and control mode offers maximum flexibility. In most separators where close control of level is desired, snap-action is preferable but may not be practical at high pressures and/or when large valves are involved. It would be unusual to use over 25% proportional control in a separator. When liquid surging is a problem, a compound control is indicated. A regular control valve is supplemented by a second one which snaps open when a surge arrives too large for the regular control. (See Chapter 11.)

The standard displacement float is satisfactory for most applications where level control is the primary goal. In the general case though its length (span) depends on the maximum liquid flow rate, vessel geometry and valve size. In a single barrel horizontal separator, for example, the level cannot be allowed to increase very much during surging. So, a large span may not be feasible.

When the vessel is designed to handle surges, the span may be estimated from desired hold-up time at maximum liquid flow rate. Hold-up time for common separations are shown in Chapter 11.

The rate of level change is found by dividing the maximum liquid flow rate by the liquid surface area in the vessel. If this exceeds about 12 cm/s [4.5 in/s] control instability could result because the level changes faster than the controller and valve can respond. In this case, vessel size may be governed by this consideration rather than gas rate or hold-up time.

By use of a longer than normal float and a high proportional setting one may control large swings in level. This may be called "flow averaging" control since the objective is to smooth out flow and not control level. This may be a less expensive, more reliable way to handle surges than a flow controller. Flow is not held constant but variations in it are dampened if the vessel and span are specified properly.

Reset is seldom required in separation systems. It may be employed sometimes in single barrel horizontal separators where relatively small level changes can affect performance.

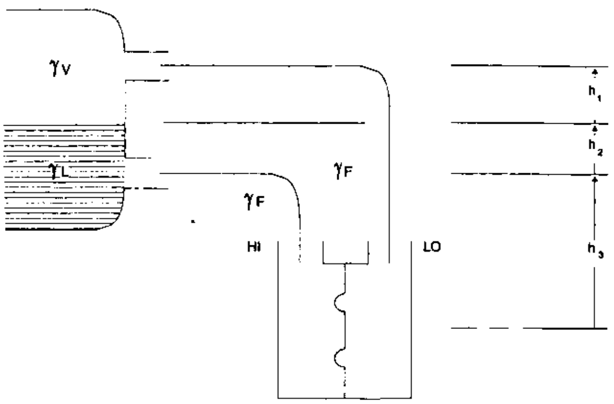
The trim size should be determined on the basis of 110% of the maximum flow to be encountered. This means that during a surging condition the valve can handle the volume and prevent the separator from filling up.

Some uncertainty always exists, of course, when one tries to estimate future flow rates. On low-pressure separators, sizing is not a serious problem because of the low pressure drop across the valve. On high-pressure gas-condensate separators, however, where the liquid flow rate is low, too large an inner valve on snap-action can blow all the liquid out of the separator and allow gas to enter the stock tank before the valve can close.

When handling very small volumes of liquid, the use of sizing curves often indicates the use of so-called "metering trim." Its use is not recommended in field applications, for the small quantities of solid materials often present may clog the valve. For this reason, nothing smaller than 0.5 cm needle trim is usually recommended in field service.

Even this may be too small unless the separation pressure is very large.

The *differential pressure (DP) cell* measures the hydrostatic head of liquid in a vessel. It is self-contained (at low pressure) or transmits a signal to a controller. A DP cell is applicable particularly in a vessel containing vaporizing or condensing liquids. The basic principle is illustrated by Equation 9.6 and the accompanying sketch. The fluid in the heads is usually non-volatile and immiscible with the fluid in the vessel. Glycol and methanol are frequently used.



$$\Delta P = (h_1 + h_2 + h_3)\,\gamma_F - h_1\,\gamma_V - h_2\,\gamma_L - h_3\,\gamma_F \qquad (9.6)$$

Where:		Metric	English
	ΔP = differential pressure	cm H_2O	in. H_2O
	h = height of fluid	cm	in.
	γ = relative density		

The self-contained unit shown in Figure 9.6 is applicable up to about 800 kPa [115 psi] within the valve sizes available. The differential head between the vessel vapor and liquid spaces is used in the pilot section to activate the valve.

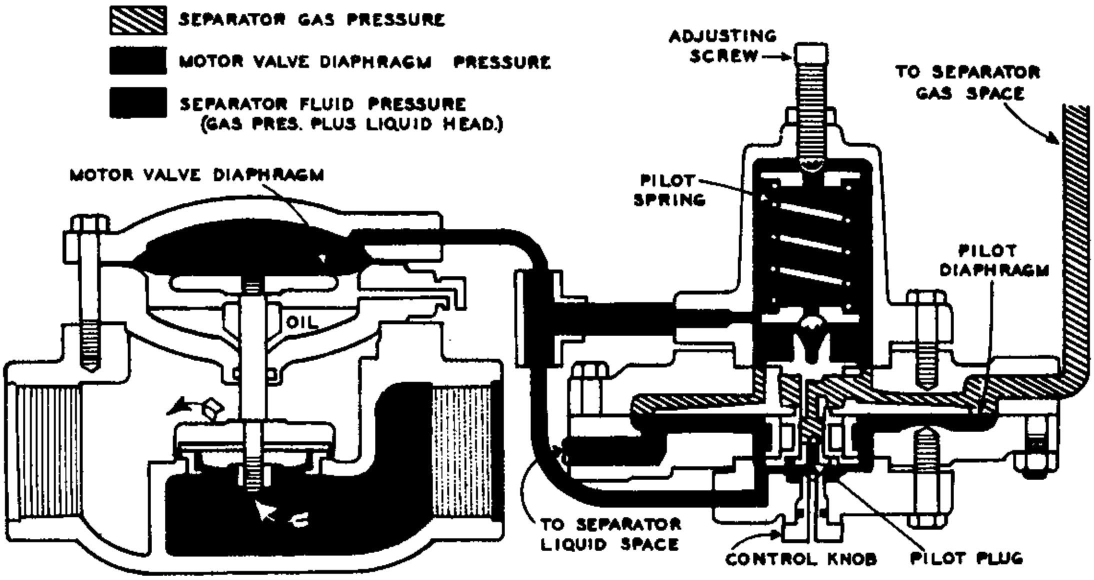


Figure 9.6 Schematic View of Low Pressure DP Controller

Other Level Devices

The above devices are the ones seen most commonly used in current production/processing operations but there are other types.

A *capacitance* instrument consists of a metal rod covered with Teflon insulation if the liquid conducts electricity. If the liquid is dielectric it consists of a metal rod inside a metal tube. The capacitance of the system varies linearly with level changes. The corresponding signal change can be used to control level.

A *sonic* and *ultrasonic* instrument uses waves above 20 kHz transmitted by a transducer/sensor head that are bounced off the liquid surface. This can be used to measure liquid-foam interfaces, solids, slurries, etc.

Radiation instruments utilize gamma rays from sources like radium or Cobalt 60. A series of measuring cells in a vertical plane detect the radiation received across the vessel. The Geiger-Mueller count rate is lower across the liquid than across the gas. Because of potential hazard problems application of this instrument are limited.

Many other measuring instruments, strain gauges and the like, also have been used, but they are of little practical value in the system discussed herein.

PRESSURE REGULATION AND CONTROL

The regulation of *back pressure* and *pressure reduction* (letdown) in a system may be conveniently divided into three categories when considering the type system needed. The low-pressure range is usually to 800 kPa [115 psig], the intermediate pressure 800-3500 kPa [115-500 psig], and the high pressure greater than 3500 kPa [500 psig]. The use of these ranges is primarily for convenience, since some types of valves can operate satisfactorily in all of them.

All pressure regulators are similar in principle, the specification of type being dependent on the process requirements, pressure drop, variation in flow rate, limitations of the loading device, and the maximum pressure. In either of these services pressure is regulated by the control of flow rate.

This flow is controlled through movement of the regulator inner valve, which is held either open or closed by some means of preloading. The amount of preloading and the size of the diaphragm used, if any, are such that the inner valve will move to the opposite extreme of travel shortly after the diaphragm pressure passes the desired working pressure. The control pressure is therefore varied by changing the amount of preload, thus upsetting the equilibrium between it and the diaphragm pressure. With most valves, increasing preload increases pressure.

The preloading may be accomplished through the use of spring compression, dead weight, or fluid pressure. Regardless of the mechanism of pressure regulation and control, there are three major purposes for its inclusion:

1. Pressure protection of equipment.
2. Control of pressure as part of the control of phase and fluid composition specifications.
3. Letdown of pressure to control the pressure or flow of a fluid to the next equipment module.

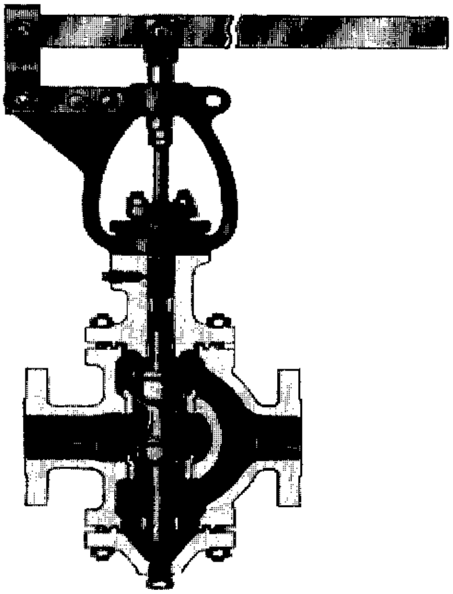
The objectives of the control system and the pressure range of the system dictate the controls chosen.

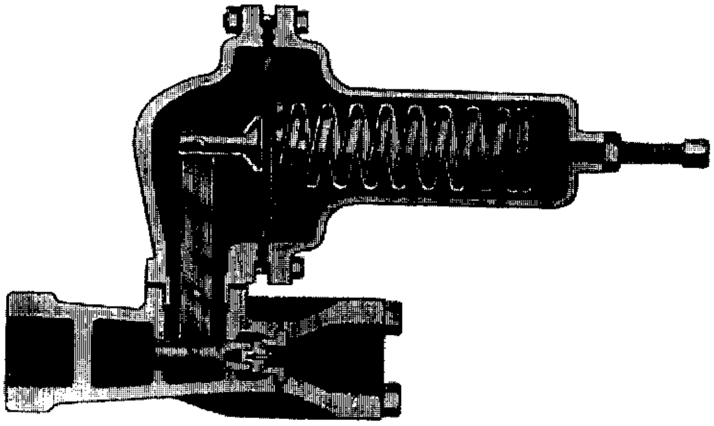
The system ultimately chosen normally falls into one of two categories:

1. Self-contained pressure regulation.
2. Pilot-operated pressure regulation.

Self-contained systems are typically used in low pressure service (less than 800 kPa [115 psia]) or in systems where small flows are required (less than 10×10^3 std m^3/d [360 Mscfd]). Two types are shown on the next page.

Since no outside connections are necessary, self-contained regulators are compact in size and give extremely dependable service. They have a simple design, are easy to maintain, and in general are more economical than pilot-actuated pressure controllers with diaphragm motor valves.





Courtesy Fisher Controls

At pressures above 800 kPa [115 psia] it usually is difficult to justify anything other than a pilot-operated diaphragm motor valve for pressure control. Non-pilot loaded valves are sometimes applicable, but usually only in those circumstances where pronounced load changes are not encountered. At these higher pressures the cost differential also becomes less, which further encourages the use of pilot-operated controls.

The top portion of Figure 9.4 is a pressure reducing regulator. Increasing downstream pressure increases diaphragm pressure on the valve causing it to start closing. The action can be changed on-site to reverse this action for back pressure control.

Pressure Controller Specification

Pressure systems involve processes whose time constants can be very small, very large or any value in between. Pressure control of liquid systems involves very small time constants due to the incompressible nature of the fluid. These systems normally display controllers with proportional and reset action using wide proportional band settings (100-200%) and relatively fast reset action (20 repeats/min).

Pressure control on gas systems usually means large time constants. As a result the proportional action should be relatively fast. One method of estimating the initial proportional band required is shown in Equation 9.7.

$$\mathrm{PB} = \frac{(200)(\mathrm{Overshoot})}{\mathrm{Span}} \tag{9.7}$$

Where: PB = proportional band, %
Overshoot = tolerable overshoot, kPa, psi, etc.
Span = transmitter range, kPa, psi, etc.

Example 9.3: Estimate the initial proportional band setting on a pressure controller with a bourdon tube range of 0-2.0 MPa. An overshoot of 150 kPa can be tolerated.

$$\mathrm{PB} = \frac{(200)(150)}{2000} = 15$$

Many pressure controllers in field service do not require reset action, inasmuch as offset is not a serious impediment to the system performance. If reset action is used it should be quite slow with typical setting of 2 repeats/min or less.

Most back-pressure regulators on field separators do not employ reset action; however, in a downstream pressure regulation, such as pressure control on a fuel-gas system, reset action may be desirable.

The control valves specified for pressure control systems typically have linear characteristics. Only where flow varies over a wide range of values, or where available pressure drop at maximum flow is less than 20% of available pressure drop at minimum flow, would an equal percentage valve characteristic be preferred.

TEMPERATURE MEASUREMENT AND CONTROL

Temperature is the most difficult of all process variables to control. Temperature control systems are characterized by large process and measurement lags. Temperature controllers (other than snap-acting) may employ derivative (rate) action to counteract the system lag and speed up controller response.

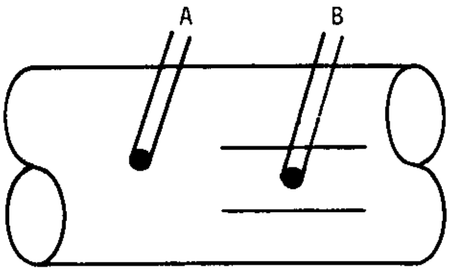
Temperature Measurement

Since temperature is a concept indicative of "molecular activity" it cannot be measured directly, but must be measured by inference. Ways in which temperature may be inferred include: expansion and contraction of liquids and metals, changes in volume or pressure of gas, change in electrical resistance and intensity of emitted radiation. The most common temperature measurement devices are:

1. Thermometers
    a. Liquid filled
    b. Bimetal
2. Filled Systems
3. Thermocouples
4. Electric resistance devices
5. Radiation pyrometers

Selection of the best sensor for a given application is a function of temperature range, accuracy, speed of response, initial cost and maintenance requirements. Each of the temperature sensing systems shown above will be discussed in light of these factors.

Regardless of the measuring device used it must be inserted into a probe to measure temperature inside a vessel or pipe.



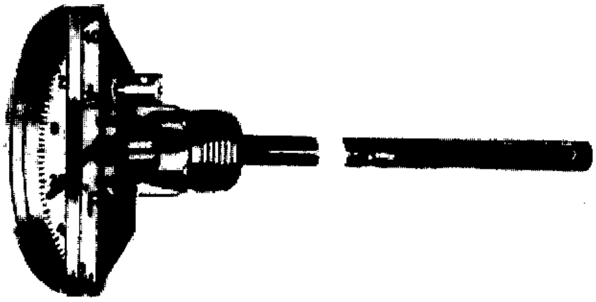
The temperature at the tip of this probe is the resultant of three heat transfer processes: heat transfer by convection between the flowing fluid and the tip, conduction along the probe, and radiation gain or loss to or from the walls. The error in the temperature measured may be particularly large in hot gas streams flowing through uninsulated pipe. Using a probe like (A), the recorded temperature may be 10-20% low. This is due to relatively poor convection transfer to the probe compared to radiant losses to the wall. Shielding the probe as in (B) prevents the tip from "seeing" the colder wall and improves temperature measurement.

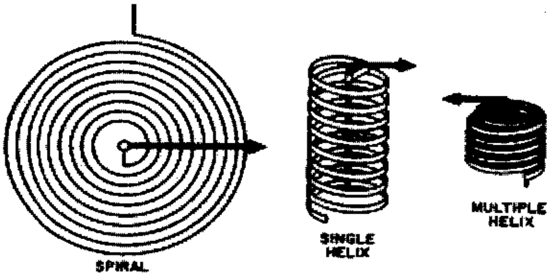
Thermometers

Thermometers used in the oil and gas industry processes are one of two types: (1) liquid or mercury filled glass thermometer or (2) bimetal thermometer.

The most common type of glass thermometer is mercury filled. It can be employed over a wide range of temperatures, °37°C to 500°C [100°F-930°F]. For temperatures below –37°C [-35°F], liquid filled thermometers employing alcohol may be used. Glass thermometers have the advantage of simplicity, low cost, and long life. The disadvantages of glass thermometers are obvious. They are easily broken, are difficult to read from a distance and because they generate no measurable signal cannot be utilized easily in a control system.

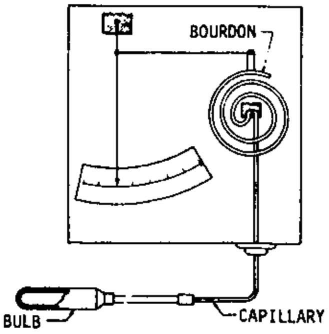
Bimetal thermometers are also widely used for temperature indication. In a bimetal thermometer, two thin strips of dissimilar metals are attached by welding, brazing, or riveting. If the metals selected have substantially different coefficients of thermal expansion, changing temperatures will cause distortion in the bimetal element which can be converted to a mechanical indication on a dial. Sensitivity is increased by using long thin bimetal elements. Examples of element types and a picture of a bimetal thermometer are shown below.(9.6)





Bimetal thermometers have the advantage of low cost, are easy to read, and are less subject to breakage. They are generally considered to be less accurate than glass thermometers, however, and need to be recalibrated periodically especially if they have been overranged or subjected to harsh operating conditions. Bimetal thermometers have been successfully used over temperature ranges of –70°C to 500°C [-94°F to 930°F].

Filled Systems



A filled system senses temperature by measuring a fluid's volume or pressure change that accompanies a temperature change. The basic system shown at right consists of a temperature sensor (bulb), a capillary tube, a bourdon tube or bellows element, and a fluid. The bulb is placed in the piping or vessel where the temperature is to be measured, and the element's motion due to a change in volume or pressure of the fluid is transmitted to an indicator or controller.

Filled systems fall in one of four classes designated by Scientific Apparatus Manufacturers Association (SAMA).

Type I	Liquid Filled Volume Change System
Type II	Vapor Filled Pressure Change System
Type III	Gas Filled Pressure Change System
Type IV	Mercury Filled Volume Change System

Various subcategories may be specified depending on the temperature range of the sensor and the type of ambient temperature compensation employed.

The effects of ambient-temperature changes on temperature measurement depend on several factors, including the type of system, its temperature range, the length of the capillary tubing and its material of construction, the fill fluid, etc. In liquid and gas systems, both the motion element and the capillary can be compensated for these effects (full compensation), or just the motion element can be compensated (case compensation). Because Type II systems are unaffected by ambient-temperature changes, they do not need compensation.

The total ambient-temperature effect on a fully compensated liquid system is usually less than ±1% of span in the range of 0 to 38°C [32 to 100°F] at the instrument or tubing when the measurement is near midrange. For a vapor system (which has no compensation), the error is less than ±0.7% of span under the same conditions.

The Class II vapor system is normally used because it is the simplest, least expensive and most responsive filled system available. The sensor output is nonlinear with temperature, but this is generally not a serious problem and actually provides higher resolution at the upper end of the scale. Fully compensated Class I systems are typically the second choice. Class III systems have the advantage of wide temperature ranges, but have the disadvantage of a large sensing bulb. Class IV systems are seldom used today.

Thermocouples

A thermocouple consists of a pair of conductors of different metals or alloys joined together at both ends. One end, the measuring junction, is placed where the temperature is to be measured. The two conductors extend out of the measurement area to the reference junction. An electromotive force (emf) is produced, and is a function of the temperature difference between the two junctions.

Because the net emf generated is a function of the temperatures of both junctions, control or compensation of the reference-junction temperature is required. This may be done by maintaining the reference junction at a constant temperature. However, this usually is not practical in a plant environment. The most common method of temperature compensation is electrical compensation which effectively cancels the voltage output of the reference junction.

Several thermocouple types are available. These are summarized in Table 9.2.

TABLE 9.2

Composition and Temperature Range of Common Thermocouples

Type	Composition	Temperature Range	
		°C	°F
B	Platinum-6% Rhodium vs. Platinum-30% Rhodium	0 to 1820	32 to 3310
R	Platinum vs. Platinum-13% Rhodium	–50 to 1768	–60 to 3210
S	Platinum vs. Platinum-10% Rhodium	–50 to 1768	–60 to 3210
J	Iron vs. Constantan	–210 to 760	–350 to 1400
K	Chromel vs. Alumel	–270 to 1372	–460 to 2500
T	Copper vs. Constantan	–270 to 400	–450 to 750
E	Chromel vs. Constantan	–270 to 1000	–450 to 1830

The major requirements of thermocouple materials are (1) they must not deteriorate; (2) they must produce a measurable, stable, electrical output; (3) they must be economical; and (4) they must be mechanically strong. Thermocouple types J & T are widely used in the oil and gas industry.

Exposed thermocouples give the fastest response time but are generally not used because of pressure, corrosion, moisture or erosion. As a result, thermocouples are almost always mounted inside a thermowell which can lengthen the response time by a factor of three to ten. Methods used to minimize the response time include insuring that the thermocouple physically contacts the thermowell wall, and filling the well with a fluid, e.g., glycol, alcohol, oil etc.

Another factor to be considered is the conduction effect. Since the thermowell extends out from the process, a temperature gradient will exist along its length, and if the tube or well is not inserted deep enough into the process, that gradient will cause inaccuracies in measurement. To eliminate this effect, the insertion length within the process should be at least ten times the sensor diameter.

Resistance Temperature Detectors (RTD's)

An RTD works on the principle that as the temperature of a material changes its electrical resistance changes and, further, that the magnitude of the resistance change is nearly a linear function of temperature. RTD's are the most accurate of the temperature measuring devices.

Materials used for RTD construction are platinum, nickel, copper and tungsten, with platinum being the most common.

The assemblies of industrial RTD's are practically identical to those of thermocouples (externally there is usually no difference in appearance). The same environmental and installation considerations apply; the same amount of attention must be paid to pressure, temperature, chemical attack, abrasion, vibration, porosity and fluid velocity; and the same types of wells are required.

Film type RTD's consist of a metallic fill (normally platinum) deposited on an insulating material with an overcoating that protects the film from the environment. It can be used to measure the surface temperature of a stationary object such as a bearing housing. Since it has little mass it responds rapidly to changes in temperature.

Radiation Pyrometers

Radiation pyrometry is the measure of temperatures, without contact, by measuring the thermal energy radiated from the object. In the oil and gas industry radiation pyrometers are used primarily for checking for hot spots on heater/boiler tubes or skin temperatures on heater walls where the location is inaccessible.

Some Temperature Control Applications

As indicated earlier, temperature control is characterized by large process and measurement lags which make control difficult. The use of derivative action to speed up controller response and cascade control to minimize the effect of load changes is very common in temperature control systems.

The most common temperature control applications in the petroleum industry are:

A. Heated fluid passes through heater.
   1. Indirect type, using:
      a) water
      b) heat-transfer oil
      c) eutectic salt
   2. Direct Type

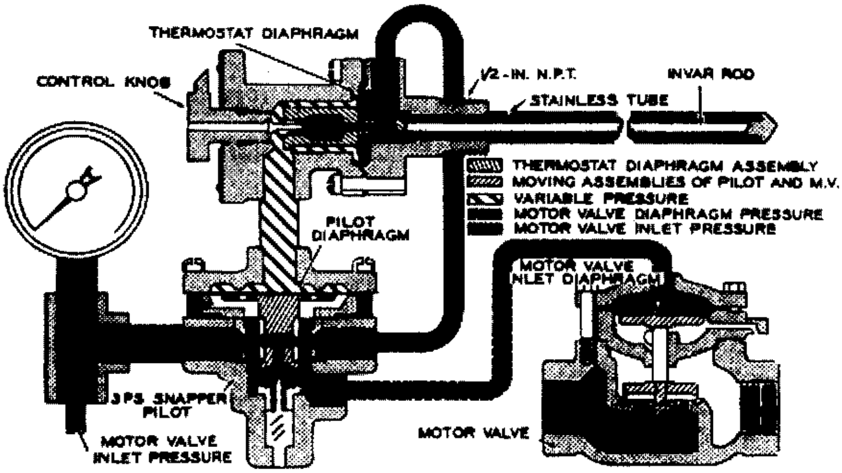
B. Heating medium circulates.
   1. Hot water
   2. Low-pressure stream
   3. Heat-transfer oil, Dowtherm, etc.

An indirect heater is basically a device whereby the burner heats a fluid bath, which in turn heats the fluid passing through the flow coils. In this manner the hot flame is isolated from the hydrocarbon, which minimizes coking and burnouts and increases the ultimate safety. The choice of fluid bath is of course a function of the temperature range desired. Indirect fired heaters are discussed in more detail in Chapter 13.

From a control standpoint, the large mass of the liquid bath used complicates the problem, particularly if accurate temperatures are required. On oil and gas lines where only approximate temperatures are needed, it is usually sufficient to control the bath temperature. However, where the heat output must vary, a controller must be placed on the flow coil, for the bath acts as a "heat sponge" and prevents rapid response to the temperature pilot.

Contrary to the usual situation, the simplest instrument is usually the most desirable for control of an indirect heater. In fact, not over 10% proportional control is recommended because the bath temperature would always lag behind the controller. This could be overcome by using an instrument with derivative response, but the cost would not be justifiable. In this application reset would not be needed in any event.

A thermostat of the type shown is very satisfactory in this service. The probe mounted in the vessel consists of an invar rod within a stainless steel tube. Changes in temperature will change the length of the tube but have very little effect on the invar rod. The relative movement of these two elements then controls the action at the pilot seats.



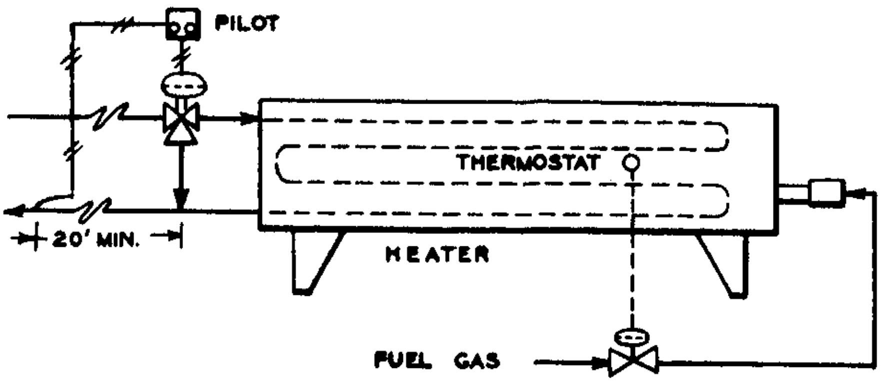
In the particular model shown, the thermostat action then actuates a snap-acting pilot, which in turn operates a simple diaphragm valve. Using this means of control, the burner is either on or off. From the control standpoint alone this is desirable, for it enables the fast transference of heat to the relative large mass represented by the liquid bath.

On large heaters though, the large surge of gas when the valve opens is sometimes felt to be undesirable and some degree of throttling is needed.

When a control is needed on the fluid through the coil, a thermostat is not entirely satisfactory, particularly if the flow rate varies widely.

The figure on the next page is an example of this type of control. The thermostat holds the bath temperature at the predetermined level, preferably greater than 10°C above the desired outlet temperature. A three-way valve then bypasses enough of the cool fluid to give the desired outlet temperature. Under design conditions some fluid should be bypassed so that control in both directions is provided.

The temperature-sensing bulb should be at least 20 ft. from the bypass point to allow mixing of the streams. This is an inherent disadvantage of the system, for it introduces an additional time lag into the system.



The three-way valve may be replaced by a less expensive two-way valve located on the bypass. The valve must be carefully sized, for the pressure drop is small through the heater coil and, consequently, it has to operate on a small pressure differential. In the final analysis, the three-way valve is usually worth the extra investment because of the flexibility it affords.

Direct-Fired Heaters

In direct-fired heaters no buffer solution is used between the fire and the flow tubes, the heat being transferred by direct flame impingement, radiation from the flame, and the hot exhaust gases. They offer the advantage of very positive heat transfer since the mass involved is relatively small. Their chief disadvantage stems from the large temperature differential that exists across the tube wall, which increases the danger of hydrocarbon coking, fouling, and ultimate failure of the tubes. It is for these reasons that they have been used largely in plants where they could be supervised and heat loads per unit volume are encountered.

The basic system usually consists of a temperature controller on the fluid outlet that controls the gas to the burner. In the event that this control fails, a high temperature shut-off is used to prevent heater damage. Proportional and reset customarily is specified for it gives good control and minimizes thermal stresses in the flow coil. The manufacturers of tubing discourage the use of on-off control, for repeated heating and cooling sets up stresses that can ultimately cause failure.

Indirect Heating Medium

Indirect heating using steam, hot oil, hot water or a heat transfer fluid such as Dowtherm is very common in process applications.

Where a considerable number of heating requirements in the same temperature range are needed, indirect heating is used. Examples include reboilers, trim heaters, feed preheaters, tank heaters, etc.

When steam is used as the heating medium the most common control system is to manipulate the steam pressure on the exchanger shell. Since the steam is condensing on the tubes manipulating steam pressure controls the exchanger driving force – hence heat transfer rate. The steam condensate is dumped from the exchanger using a steam trap.

An alternative method is to manipulate the level of steam condensate trapped in the exchanger thereby controlling effective transfer area. This method requires an oversized heat exchanger to produce a flexible control range.

When hot oil or a heat transfer fluid is used as the heating medium the standard control practice is to manipulate the rate of hot fluid through the exchanger. The pilot controls the process fluid temperature by manipulating the hot fluid rate.

Both of these systems seldom give satisfactory performance unless the temperature controller is cascaded with another controller handling the load changes on the manipulated variable. This is discussed in the next section.

In estimating controller setting for a temperature control systems, Equation 9.7 can be used for the proportional band. Reset action should be relatively slow, around 0.2 repeats/min. Rate action can be estimated by setting it equal to the time constant of the thermowell. The derivative action then cancels the measurement lag introduced into the system by the thermowell.

FLOW MEASUREMENT AND CONTROL

Flow systems are characterized by very rapid responses for fluids encountered in hydrocarbon systems. When the flowmeter is located reasonably near the control valve, the flow process will respond faster than the instruments in the control loop. Because of the negligible lags in the flow system and the very noisy signal received from the flowmeter, the proportional band setting on a flow controller is typically very wide (200%). Reset action is rapid (approx. 20 repeats/min).(9.6) Rate action should not be specified on a flow controller because it responds to the noise in the system and amplifies it, leading to unstable control.

Flow controllers are seldom used alone. They almost always are cascaded with another control loop, e.g., pressure, level and temperature. There are two reasons for this. First, flow controllers will control flow without regard for any other variable in the system. For example, a flow controller on the outlet of a tank may drain the tank dry or allow it to run over unless reset by a level controller. Secondly, in any control system, whether temperature, pressure, etc., the valve is manipulating the flow rate of a fluid. A flow controller can respond rapidly to load changes on the manipulated variable preventing those changes from affecting the controlled variable.

Flow Measurement

Flow is the most difficult of the process variables to measure. It cannot be measured directly as can pressure and level, rather it is measure by inference ... differential pressure, turbine rotation, etc. In addition, flow measurement must be made in the fluid itself, not in parallel to the fluid. For such reasons a great many flow measurement techniques have been developed. This text will concentrate on those used in the oil and gas industry.

The choice of a meter is determined by its mechanical and measurement reliability and its range. The reliability is a matter of experience and testing. The range needed must be established by preliminary design calculations.

Measurement of flow rate traditionally has been purely mechanical or has used some type of differential pressure measurement. The orifice meter has been the most common type. Now the designer has a large array of meters available for choice. Some of these possess charisma but are not suitable in a given environment.

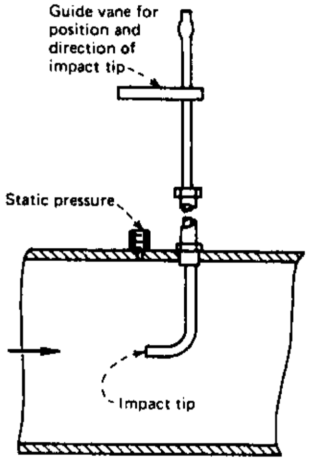
Simplicity of the meter is an important consideration. This is particularly important in isolated areas or where the availability of skilled technicians is limited (or nonexistent). A meter suitable for a large plant may not be suitable for a production installation.

In assessing a meter consider the difference between accuracy and *precision.* Both are errors but of a different type. Accuracy is the relationship between the true rate and that measured. Some "proving" procedure is used to determine the amount of error. Precision is the repeatability of a measurement – right or wrong. An instrument may be very precise and yet be very inaccurate.

In the section which follows I will review some of the meters available and then discuss a few pertinent ones in more detail.

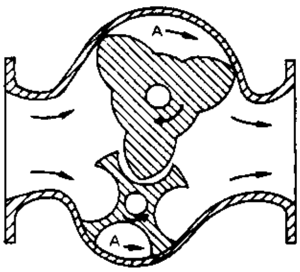
Pitot (Impact) Tube

The static pressure is measured at the wall and the dynamic (flowing) pressure is measured by the tip. It is suitable for some measurements but the accuracy is not suitable for custody metering. The greatest error is in measurement of static pressure. The tip also may become fouled with liquids or solids.



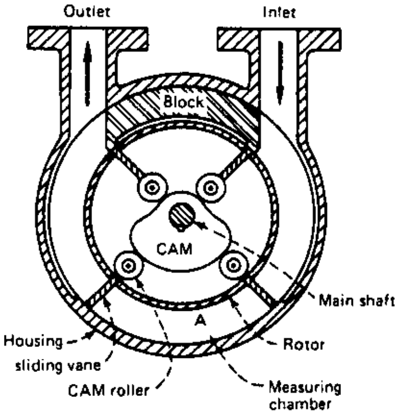
Rotary Meter

This is used primarily for measuring custody transfer of liquids. Its rangeability is about 5:1 and is very accurate. It also offers a low pressure drop. Capacity is limited by rpm. Also, the rotating element is difficult to replace.



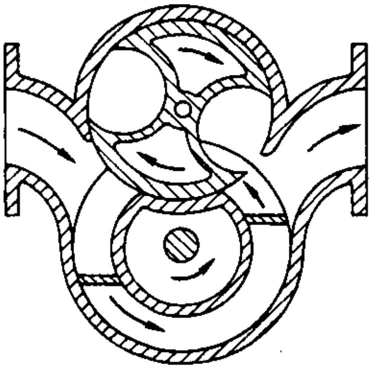
Sliding Vane Meter

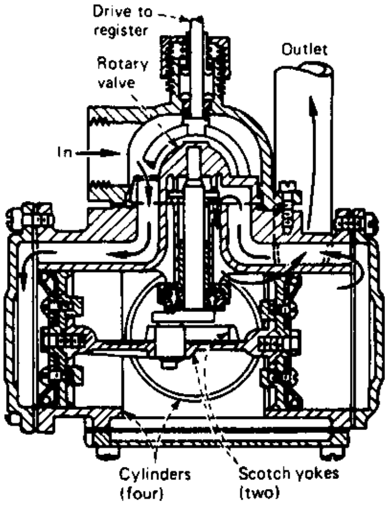
This is equivalent to the rotary meter and shares its advantages and disadvantages. It is used for the custody transfer of gasses as well as liquids.



Rotating Vane Meter

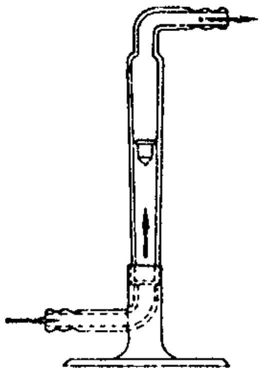
Both the rotor and vanes rotate. It has a rangeability of 5:1 and offers excellent repeatability. It is used for custody transfer of both gasses and liquids.





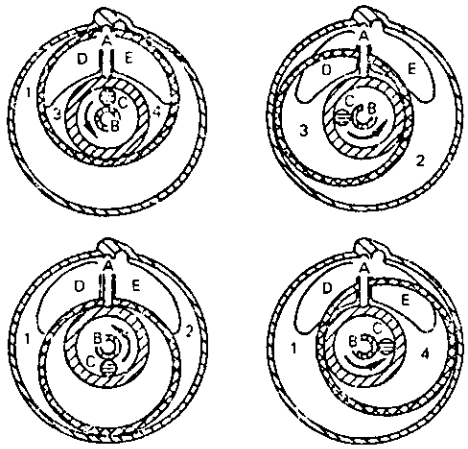
Reciprocating Piston Meter

In this meter, opposing pistons connected to a central crankshaft alternately fill and empty. Its capacity is limited but is the most accurate of the positive displacement meters. Its range-ability is 50:1 and is usually used for metering low flow rate chemical streams.



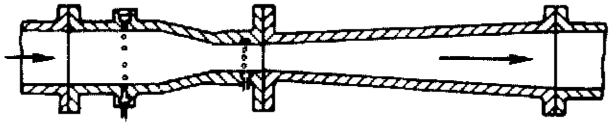
Rotameter

This is a variable orifice meter useful for the indication of flow rates in operations. It is economical in sizes up to 10 cm [4 in.], at moderate pressures. It can transmit electrical or pneumatic signals. Its rangeability is about 10:1.



Rotary Piston Meters

An eccentric cylinder rotates within the case. The volume inside and outside this cylinder alternately fills and empties. It is used primarily for accounting purposes on viscous liq-uids.

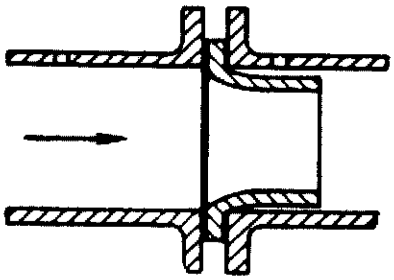


Venturi Meter

This is a variation of that class of meters whereby flow is measured indirectly through the first law of thermodynamics by measuring a pressure differential. The analysis is equivalent to that of the orifice meter.

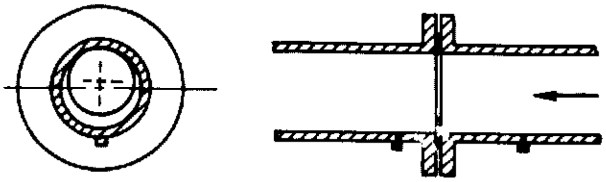
It offers low pressure loss for measurement of all kinds of fluids. It has a rangeability of about 3.5:1.

Flow Nozzles



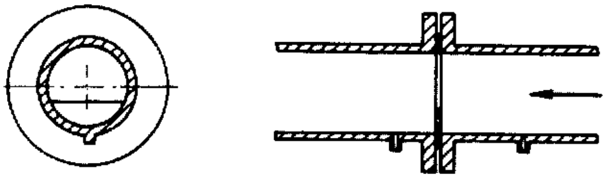
These are used for high velocity streams and will pass about 60% more flow than an orifice plate for the same line size at the same pressure differential. By virtue of a rounded edge it does not "cut out" as readily when solids are present in the fluid. The rangeability is about 3.5:1.

Eccentric Orifices



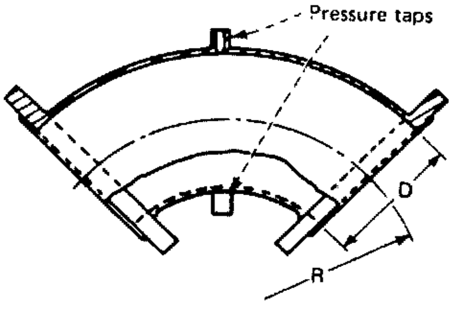
These are used for many two-phase streams including those containing solids. Substantial straight-line pipe sections are required both upstream and downstream.

Segmental Orifices



These possess the same basic applicability as the eccentric orifices with an additional advantage that they do not dam solids on the upstream side of the plate. These are used primarily for large line sizes and low fluid viscosities.

Centrifugal (Elbow) Meters



This is used for large pipe sizes primarily. It is based on the centrifugal force induced as the fluid changes direction. This centrifugal force is a function of velocity, R and D, as well as fluid properties. A common pipe elbow is commonly used. If very accurate results are desired calibration is necessary. Its rangeability is about 3:1.

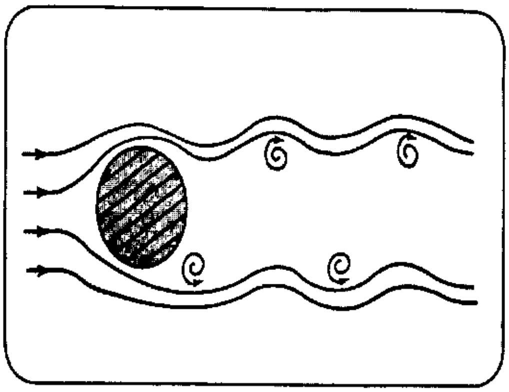
All of the above are used in selective applications. The largest volume of metering is done by turbine and orifice meters. Sonic and vortex meter usage has continued to increase but they have enjoyed only limited employment in production operations.

Vortex Shedding

This is a newer meter for measuring all fluids that do not have too high a viscosity. It has no moving parts, wide rangeability and is linear over wide ranges of flow.

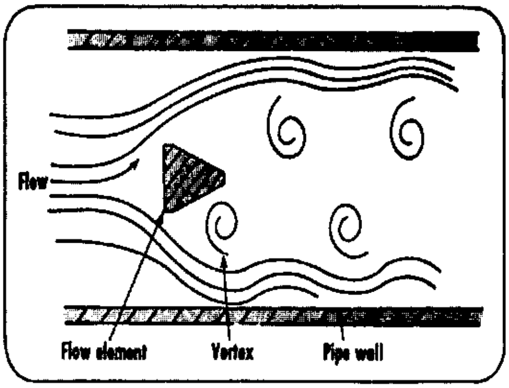
The figure at the top of page 287 shows the principle involved when a fluid flows around a blunt object. The flow is unable to follow the surface of the object and sheds from it at some point to form a continuous series of eddy currents. The frequency of shedding is proportional to flow velocity and inversely proportional to the diameter of the object.

This shedding used to be sensed by thermistors or by what was called a "shuttle ball" contained in the fixed flow element. Movement of this ball was detected by a magnetic sensor. But this system was plagued with plugging and mechanical failures. The thermistor type of sensors exhibited unsatisfactory signal-to-



Karman Vortex Trail

(Behind a circular cylinder - idealized)



Vortex Flow Pattern

OGJ

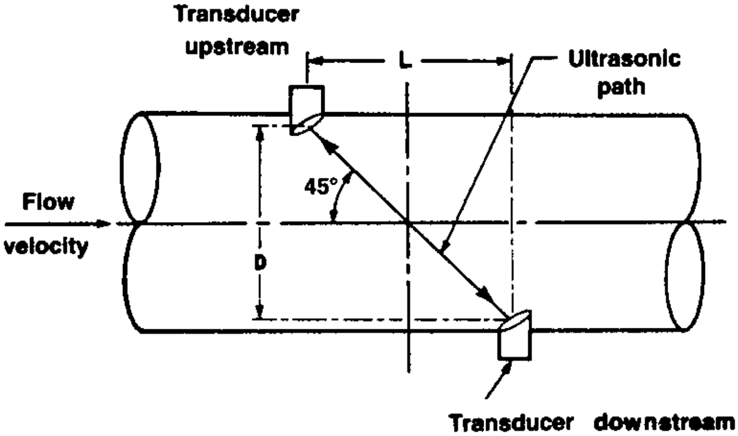
noise ratios, were easily fouled and sensitive to vibration. Piezoelectric sensors have eliminated the worst of these problems although vortex meters are not recommended in the vicinity of rotating machinery or vibration of any magnitude.(9.7, 9.8)

Meter accuracy should be about ±1% with a rangeability of 15:1. The Reynolds number should be at least 10 000 to achieve the above accuracy.

The vortex meter is a suitable alternative to a turbine meter when handling fluids that may cause turbine lubrication problems, like water, steam, alcohols, acids, etc.

Sonic Meters

Several types of sonic type meters are produced, the doppler and contrapropagatings. The latter is applicable to natural gas metering.(9.9) This type of meter offers promise for large gas lines because it is not in the gas pipe proper, causes no pressure loss, and is portable. It can measure flow in either direction and indicate the direction. The rangeability is about 50:1.

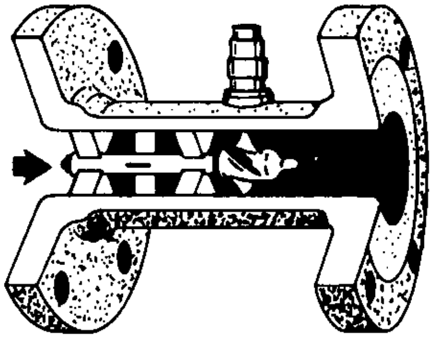


As shown, two transducers are used which alternately transmit and receive sonic pulses. Flow is obtained by measuring the time difference between the two waves over the same path, one with and against the flowing fluid.

In tests on large pipelines the sonic meter agreed with orifice measurements within about ±0.5% for an accuracy of ±2.0%.

Turbine Meters

The turbine meter has a wide range of applicability except for high viscosity fluids. Rangeability is about 20:1 for liquids and up to 100:1 for gases in some instances.



Turbine meters come in various configurations, but they all transmit a signal based on rotation speed which in turn is a function of flow rate and fluid properties. It is difficult to generalize about this type of meter for some have proven very good and some very bad in a given application. For this reason, the choice should be based on experience with a given model in a given service.

Bearings can be troublesome. Ball bearings may be susceptible to damage by abrasive solids; sleeve bearings tend to be more troublefree but rangeability may be reduced.

A screen is often desirable. If used, it should be strong enough to stand the pressure if becoming plugged. In a liquid meter, flashing, cavitation or too much vapor should be avoided since vapor can overspeed the turbine and cause bearing failure. If the line is to be heat traced it is best not to mount the signal preamplifier on the meter body to prevent overheating.

When properly selected, turbine meters are very acceptable. Their acceptability for custody transfer has been slow in some areas due to lack of proven experience compared to the orifice meter. Liquid turbine meters have been accepted more readily than those for gas service because they could be tested more readily.

Repeatability is often ±0.5% on test and accuracy is within ±1%. Like all meters these values are a function of meter condition.

The axial flow turbine meter has proven useful for high gas flow rates. Its rangeability is up to 100:1 and it has a linear response to flow rate. A summary of performance of these meters is found in References 9.10-912.

ORIFICE METERS

The orifice meter is the most common custody transfer instrument for natural gas. For many years it has been used based on calculations made with an empirical flow equation from AGA Report No. 3 or ANSI/API 2530. The equation following and the tables in Appendix 9B are based on these standards.

ISO 5167 has been published which is a new standard not compatible completely with the traditional standards. The result is a continuing study to improve on the correlation and eliminate some of the anomalies that are now known to exist. It has been found that fluid behavior in meters is very complex and traditional empiricism may not be adequate. Certainly it has been found that this meter is not as accurate as once thought.

The traditional basic equation is:

$$q_h = C' \sqrt{h_w P_f} \tag{9.8}$$

Where:
q_h = std cu ft/h at base conditions
C' = $F_b \times F_{pb} \times F_{tb} \times F_g \times F_{tf} \times F_r \times Y \times F_{pv}$ (for gas)
h_w = differential pressure, in. of water
P_f = static pressure, psia

The following values of C′ apply for steam and liquid orifice meters:

Steam: $C' = F_b \times F_s \times F_r \times Y$ Liquid: $C' = F_b \times F_{gt} \times F_r$

Appendix 9B at the end of this chapter outlines the "F" factors in English units.

Until such time as standards are published in metric units, the best approach probably is to convert to English units and use the tables. For the metric units used in this book (15°C and 100 kPa) the following conversion factors are convenient.

q_h : std m^3/h x 34.92 = std cu ft/h, std cu ft/h x 0.0286 = std m^3h

h_w : cm H_2O x 0.394 = in. H_2O , kPa x 4.02 = in. H_2O
in. H_2O x 2.54 = cm H_2O , in. H_2O x 0.249 = kPa

P_f : MPa x 145 = psia , kPa x 0.145 = psia; psia x 6.89 = kPa

F_{pb} = 1.0156 for pressure base of 100 kPa
= 1.0023 for pressure base of 14.7 psia

F_{tb} = 0.9981 for a base temperature of 15°C (59°F)

F_{tf} : Convert to °F ; (°C x 1.8) + 32 = °F

$F_r = 1 + (b/\sqrt{h_w P_f})$, where "b" is found from Appendix 9B

$F_{pv} = \sqrt{1/z}$

$F_g = \sqrt{1/\gamma}$ @ base conditions

In custody transfer there are special considerations not summarized herein. These include meter factors for mercury and bellows meters, use of square root meters, standardization of measurements, etc.

Obtaining good metering with an orifice involves the following considerations:

1. The meter range should be such that the differential pressure may be read in the middle 60-70% of the recorder scale.
2. The static tube should have a rating at least 125% of the maximum anticipated pressure, but no more than 300-400%.
3. Mechanical bellows (dry) recorders are preferable over the mercury type, even though they have a higher first cost, because of lower operating and maintenance costs.
4. Periodic calibration of the meter run should be conducted with a frequency and accuracy compatible with the desired performance.
5. With an orifice meter, the plates should be regularly inspected and replaced if worn or pitted. Using poor plates is false economy.
6. Surges of gas and liquid must be minimized. Separation ahead of the meter might be necessary.
7. Oftentimes a processing unit is fed by a number of gathering lines and sells residue gas to a single line. Both purchases and sales meters are required. If feasible, all purchase meters should be at the plant instead of on the lease. The sum of the amount metered in will not equal that metered out, even after correction for plant usage and shrinkage. The inlet meters can best be used to prorate the amount of gas sold back to each individual gathering line, using the outlet meter as the absolute quantity.
8. Provision for temperature measurement and gas sampling should be made at each meter run. The sample connection should be separate from the meter piping.

9. Always be sure the chart range on the recorder checks that of the meter with regard to both pressures and time. Be sure the proper orifice size is noted. These are obvious but are not done in a surprising number of cases.
10. Calibrate the meter run so that the actual I.D. of pipe used is known. The published wall thickness for seamless pipe may vary by as much as 12.5% for a given O.D.
11. The diameter of orifice used should be such that the ratio of its diameter to that of the pipe I.D. B will fall between the following limits:

Pipe taps: 0.20 – 0.67 Flange taps: 0.15 – 0.70

The main problem in good orifice metering is poor attention to detail. "Beat up" orifice plates, improper meter runs, and poorly maintained meters all introduce error by default.

Example 9.4: 2.0 x 10^6 std m^3/d of a 0.65 relative density natural gas are to be metered at 27°C and 6.5 MPa in an orifice meter using flange taps. At this condition the compressibility factor, z, equals 0.89. What size orifice plate and meter run should be used? Assume $h_w = 50$ in. H_2O.

$$q_h = \frac{(2.0 \times 10^6)(34.92)}{24} = 2\,910\,000 \text{ scf/h}$$

$$C' = \frac{q_h}{\sqrt{h_w P_f}} = \frac{2\,910\,000}{\sqrt{(50)(6.5)(145)}} = 13\,410$$

$$F_{pb} = \underline{1.0156}\ , \quad F_g = \sqrt{1/0.65} = \underline{1.24}$$

$$\text{Temp. of } 27°\text{C} = 80.6°\text{F}\ , \quad F_{tf} = \underline{0.9808}$$

$$F_{pv} = \sqrt{1/0.89} = 1.06\ , \quad F_{tb} = 0.9981$$

Ignore F_r and Y temporarily because they depend on pipe and orifice size. Solve for F_b.

$$F_b = \frac{C'}{F_{pb} \times F_g \times F_{tf} \times F_{pv} \times F_{tb}} = \frac{13\,410}{1.0156 \times 1.24 \times 0.9808 \times 1.06 \times 0.9981} = \underline{10\,260}$$

Several combinations of orifice to pipe ratios will fall within the recommended β ratio.

Pipe	Orifice	β
10"	6.75 in.	0.675
12"	7.0 in.	0.58
16"	7.0 in.	0.44

The larger the run, the lower β, but the higher the cost. Once a size has been chosen the values of F_r and Y could be calculated but their effect on the choice is trivial.

Heating Value

Custody transfer by heating value is becoming more common. This is a measure of gas quality. It also is convenient for it can be done independent of the pressure or temperature base conditions.(9.13, 9.14)

The *gross heating value* is the energy evolved on burning a standard volume of gas when the total combustion gases are cooled to the reference temperature and the water of combustion condensed. The condensing water releases its latent heat which is theoretically available for heating something. This represents the maximum possible heat available, thus the word "gross."

The *net heating value* is a lower number based on the assumption that the water in the combustion gases remains in the vapor state. This is the usual circumstance in most heating equipment.

Heating value may be measured with a properly calibrated calorimeter or calculated. The latter assumes ideal gases which will potentially introduce an error of about 0.2% – one cu. ft. of actual gas will ordinarily contain about 1.002 cu. ft. of ideal gas. This error is trivial for most producing and processing operations. Calculation is made by summation of the mole fraction of each component times its heating value, in the same manner as one may find pseudocritical conditions. The following gross heating values can be used in the computation:

Component	GHV per Unit Volume of Dry Gas @	
	60°F	14.7 psia
Methane	1010 Btu/ft^3	39.28 MJ/m^3
Ethane	1769	68.79
Propane	2518	97.92
iso-Butane	3256	126.6
n-Butane	3264	126.9
iso-Pentane	4001	155.6
n-Pentane	4009	155.9
Hexane	4756	184.9
Heptane	5503	214.0

Unfortunately, calculation is sensitive to accuracy of gas analysis which might easily involve an error of several percent.

Wobbe Index

This is a number used to compare fuel quality for burner use. Two fuels with the same Wobbe number (within ±5%) will deliver about the same heat at the burner head and are reasonably interchangeable. It is defined as:

$$\text{Wobbe No.} = \frac{\text{GHV}}{\gamma^{0.5}}$$

It ordinarily is quoted without units. But remember, it will have a different value when GHV is expressed in MJ/m^3 instead of Btu/ft^3.

Two-Phase Flow Measurement

The presence of liquid in a gas stream is never a desirable metering situation. There are instances, however, where it is inevitable, particularly in field operations. It has been found that one may use an orifice meter to measure two-phase flow, providing the flow rate of both phases is steady enough to yield stable output information. Murdock (9.15) presented the basic technique. Palm *et al.* show its application for determining steam quality.(9.16) This technique is limited by several factors. The volumetric gas-liquid ratio should exceed 100:1. Also, the ratio mass flow rate (lb per hr per sq ft) to pressure (psia) should exceed 100. This latter criterion will determine proper pipe size in the meter run. Murdock's correlation is for pressures to 6.4 MPa [929 psia] water condensate systems.

ALARM AND SHUTDOWN SYSTEMS

Alarm and shutdown systems are used for three reasons:

1. Protection of equipment
2. Protection of personnel
3. Orderly operation in an unattended environment.

The complexity of the systems varies with the application. An unattended lease facility may require only a few shutdowns to protect from equipment failure while an offshore platform may have a quite sophisticated system because of its location and the potential for loss of both lives and equipment.

The desire for a safe system is proper. However, in some instances, this has become such a fetish that company and government regulations have led to a massive "overkill" of the problem. Frequently, the safety system is as likely to fail as the system it presumably protects. A more common problem is that the safety system prevents normal operation of the process as loads vary.

A brief review of the available instrumentation is discussed below. Alarm and shutdown systems for particular processes or equipment will be discussed in the appropriate chapter.

Pressure Switches

Indicating or non-indicating devices are used. The scale on the indicating types often are not very accurate. Non-indicating types are not intended for operator adjustment. Usually, they are equipped with a locknut requiring screwdriver adjustment by a qualified technician.

Some switches have too little differential range – the span between actuating and resetting when near the set point. There is a minimum span in a given operation. Attempting to operate with less than this is one of the operational difficulties often encountered. If the switch cannot reset itself to clear the alarm after return to normal conditions, a kind of chaos can result. As with all controls, the differential span desired is a function of the range of the switch.

Shown on the following page are four types of pressure switches commonly employed.

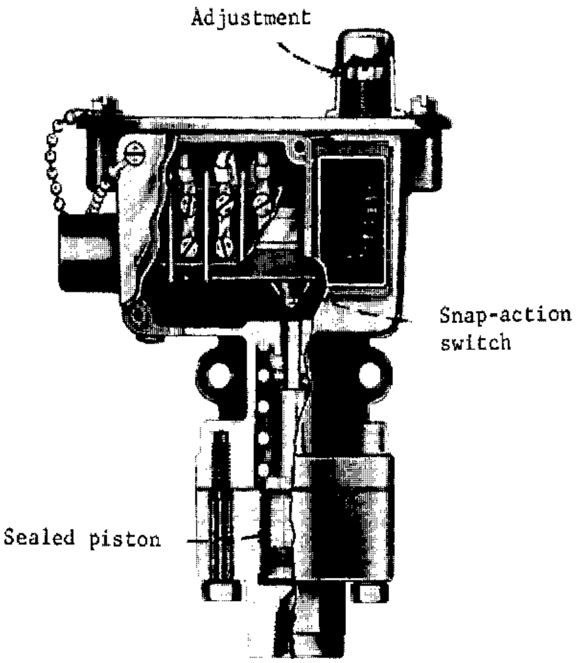
There must be a sizable range in pressure between normal and alarm conditions. Otherwise alarm is indicated during normal pressure fluctuations. For example, with a low head pump the shutdown pressure may be only slightly above nominal operating pressure and too many alarms (or shutdowns) may occur. In a case like this a flow switch rather than a pressure switch may be the best choice. As I continue to emphasize for all controls, we must choose a device that is compatible with process characteristics.

Flow Switches

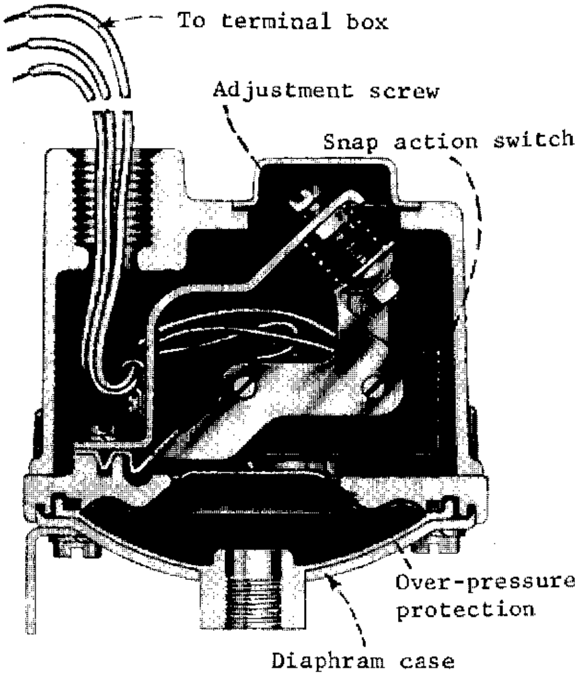
One type of simple flow switch is a velocity sensing device that has a vane or paddle located in the flow stream. These cannot provide precision but are very suitable in streams where detection of minimum flow is the primary criterion, such as low flow shutdown on a pump.

The rotameter in combination with a switch is an excellent choice for this application. However, there is a limit to the capacity available. The flow span is about 10:1.

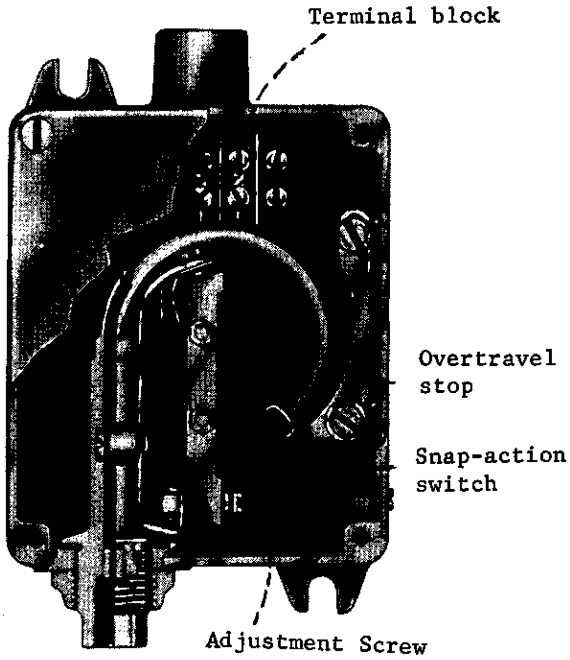
The orifice meter is capable of triggering an alarm and shutdown through transmission of a differential pressure measurement. However, orifice meters are quite expensive and there is also a problem of rangeability. It is a square root instrument. As a result, the lower 25% of the flow range represents only 6.25% of instrument output range. Resolution and sensitivity are too poor in this area to permit use in an alarm or shutdown system. In low flow applications it is recommended that alarm-point flow rate be no less than 10% of the instrument output range – or about 32% of the flow range.



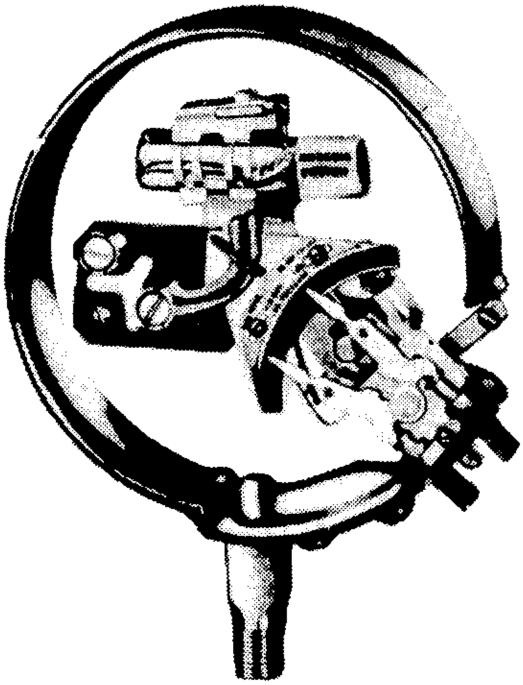
Piston Type



Diaphragm Type



Bourdon Type



Bourdon Tube

Temperature Switches

There are three basic types. A filled system either Class I or II, connected to a pressure switch through a capillary tube. Internal pressure is calibrated against temperature as in a normal temperature controller.

A vapor pressure switch is similar.

Bimetallic switches are compact and very flexible but must be mounted in the process stream. They are subject to corrosion and plagued with all the ills that befall a device in a "dirty" stream. They may be less than satisfactory when vibration is present.

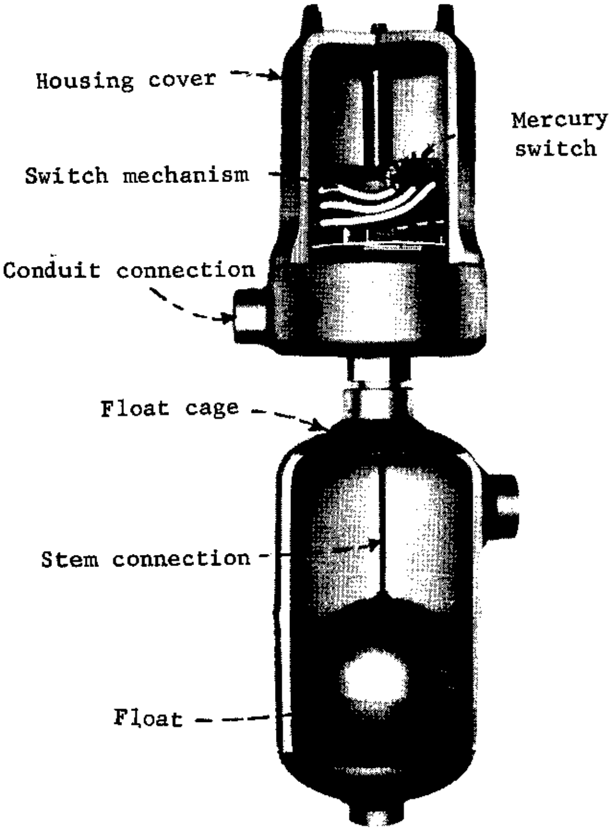
A thermocouple is reliable in itself but it requires accessory equipment that may not be suitable under adverse conditions. The small differential voltage involved must be measured by a sensitive meter relay or by an electronic amplifier with a system to activate the alarm. It is indeed a very sensitive system unless properly protected and maintained.

Level Switches

The most common float type is shown on the right. The float is directly coupled to a switch. The float can be inside the process vessel or outside in its own chamber. Both have their inherent advantages. The outside float is away from process disturbances, but one must be sure that debris in the bottom of the float cage does not hinder float movement or that any liquid entering does not hydrate or freeze.

Conductivity or capacitance probes have no moving parts and may be a satisfactory choice with corrosive fluids or at very high pressures. Obviously, the conductivity probe has no application with non-conductive fluids because it measures the conductivity between the probe and the vessel wall, or between the two probes. The capacitance unit measures the change of capacitance between the probe and the vessel wall. An insulated probe is used for conductive fluids and a bare probe for non-conductive fluids.

There are other more "exotic" types – nuclear, ultrasonic, photo cells, light refraction and the like. They are expensive and sensitive but often require technician capability not readily available. I tend to favor simple, more rugged instruments when suitable.



Float Type Level Switch

Vibration Switches

The simplest use a mass inside the switch. When vibration exceeds the set limit it overcomes the inertia of this mass to activate the switch. This type is not very sensitive or very precise but requires no external power source.

An electronic system may be used to measure radial shaft position relative to the bearing housing or shaft axial position to measure thrust. They can detect vibration less than 0.000 25 cm. Two functions can be incorporated – alarm followed by shutdown.

PRESSURE RELIEF SYSTEMS

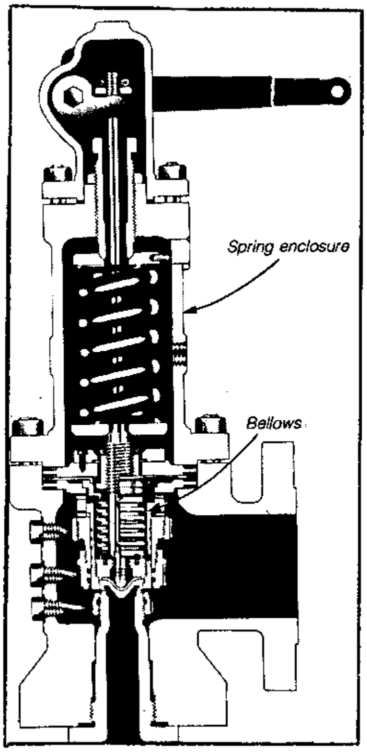
The design, selection and installation of pressure relief systems is discussed in many codes. Those listed below are widely used in the oil and gas industry.

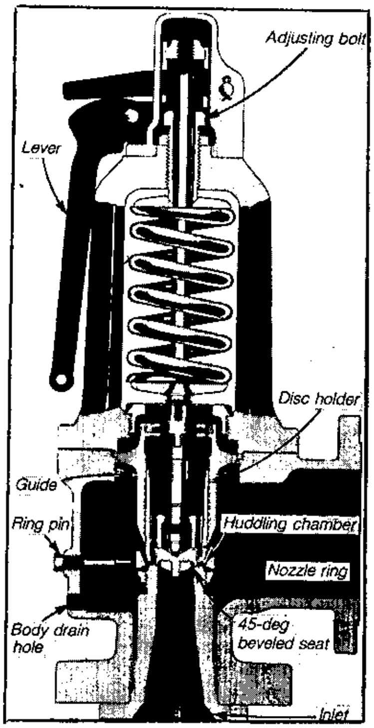
1. Section VIII ASME Boiler and Pressure Vessel Code
2. API RP520 Recommended Practice for the Design and Installation of Pressure-Relieving Systems in Refineries: Part I Design Part II Installation
3. API RP521 Guide for Pressure Relief and Depressuring Systems

Pressure relief devices may be placed in four categories:(9.19)

1. *Relief Valve* – an automatic pressure-relieving device actuated by upstream pressure that opens in proportion to pressure increase above set pressure.
2. *Safety Valve* – The same as (1) except valve possess rapid full opening or "pop" action.
3. *Safety-Relief Valves* – Suitable for either safety or relief depending on application.
4. *Rupture Disk* – a thin metal diaphragm held between flanges.

Two different types of safety valves are shown following:





Many internal designs are used to prevent leakage near the opening point, particularly after the valve has popped previously. Relief valves can be very sensitive to inlet and outlet line lengths.(9.19, 9.20)

A rupture disk is often the last resort safetywise. It may be set to burst slightly higher than a valve which serves as an early warning. It also may be used in series with a valve to prevent leakage and corrosion.(9.21)

Rupture disks have the advantage of low cost, simplicity, no breakage and give an extremely fast response. However, they are subject to stress fatigue, cannot be tested while in service, and when ruptured the entire system is depressured and must be shut down. Rupture disks are normally confined to low pressure lease operations.

Sizing

The following equation may be used for sizing safety relief valves in vapor services.

$$w = (B)\, CK_0AP_1 \sqrt{M/ZT} \qquad (9.9)$$

			Metric	English
Where:	w =	weight flow through valve	kg/hr	lb/hr
	B =	conversion constant	7.6×10^{-5}	1.0
	C =	factor depending on specific heat ratio	–	–
	K_0 =	valve discharge coefficient	obtain from manufacturer	
	A =	effective discharge area of valve	mm^2	in^2
	P_1 =	upstream relieving pressure	kPa(a)	psia
	M =	gas molecular weight	–	–
	T =	inlet temperature	K	°R
	Z =	compressibility factor	–	–

The upstream relieving pressure, P_1, is equal to the set pressure plus the allowable overpressure and atmospheric pressure. Overpressure is defined as the pressure increase over the set pressure of the relieving device. If the set pressure is equal to the maximum allowable working pressure (MAWP) of the vessel, the overpressure is 10% of the MAWP for unfired pressure vessels. If set pressure is less than the vessel MAWP the overpressure may be greater.

Equation 9.8 is to be used when flow through the valve is at critical flow which is typically the case if the pressure across the ratio is greater than 2:1. If the backpressure on the valve exceeds the critical flow backpressure, the equation must be modified to account for downstream pressure.

Relief valve manufacturers use standard orifice sizes in relief valves. These standard sizes are summarized in the table following along with the values of "C."

Orifice	Area	
	mm^2	in^2
D	71.0	0.110
E	126.5	0.196
F	198.1	0.307
G	324.6	0.503
H	506.5	0.785
J	830.3	1.287
K	1186.	1.838
L	1841.	2.853
M	2323.	3.60
N	2800.	4.34
P	4116.	6.38
Q	7129.	11.05
R	10 323.	16.0
T	16 774.	26.0

k	C
1.00	315
1.10	327
1.16	333
1.20	337
1.25	342
1.30	347
1.35	351
1.40	356

Where: k = heat capacity ratio = C_P/C_V

Installation

Sound engineering practices for the installation of relief valves can be found in the codes previously specified.

Important factors to consider are the size and orientation of the discharge piping. The discharge pipe size should never be less than the inlet pipe size and in many cases is at least one pipe size larger. Backpressure induced by piping resistance should not be more than 2 to 3% of set pressure or the relief valve may chatter. If the relief valve is to be blocked in for service, a full opening block valve should be used and when in service, the block valve should be locked open.

When a relief valve opens, a blowout thrust is imposed on the centerline of the outlet pipe which induces a bending moment on the vessel or piping. It is often necessary to reinforce the piping or vessel wall or to support the relief valve to minimize this effect.

Finally, a relief valve review maintenance program is absolutely necessary to insure proper operation and response of the valve when it is required.

NOISE CONTROL

I regard noise control as a sound operating practice, notwithstanding legal restrictions. Excess noise causes fatigue and leads to less alertness and more inefficient operations. Different government agencies in many parts of the world have adopted standards which are generally compatible with the summary which follows.

The level of sound is measured on a relative logarithmic scale which measures the pressure of the sound energy absorbed by the ear compared to a reference pressure level. For most current applications this reference pressure is 0.0002 microbars. The unit of the scale is the decibel (dB) which is related to pressure by the equation

$$\text{dB} = 20 \log (P_1/P_o) \tag{9.10}$$

Where: P_1 = pressure of the sound being measured
P_o = reference pressure

The table on the following page shows decibel levels of some common noises. The response of persons varies. A level that may bother one person may not affect another. A person can develop some tolerance for a sound heard repeatedly.

Frequency of the sound is important. A healthy human ear can normally detect sounds in the range of 20-10 000 hertz (cycles per second) although it might not be conscious of sound above 10 000 hertz. We are seldom exposed to pure tones in our audio range since most sounds are a mixture of frequencies. Consequently the audio range has been divided into a series of eight frequency bands called "octave bands."

The sensitivity of the human ear depends both on frequency and intensity (pressure). Sounds of equal sound pressure level may not sound equally loud. At 100 dB, for example, frequencies between 20 and 1000 hertz may sound equally loud. At lower pressure levels, lower frequencies do not seem as loud as the 1000 hertz tone. At low pressure levels the human ear is more sensitive to high frequency than low frequency.

The use of octave band analysis is somewhat complicated so a common method uses an A-B-C weighting network of sound level meters.

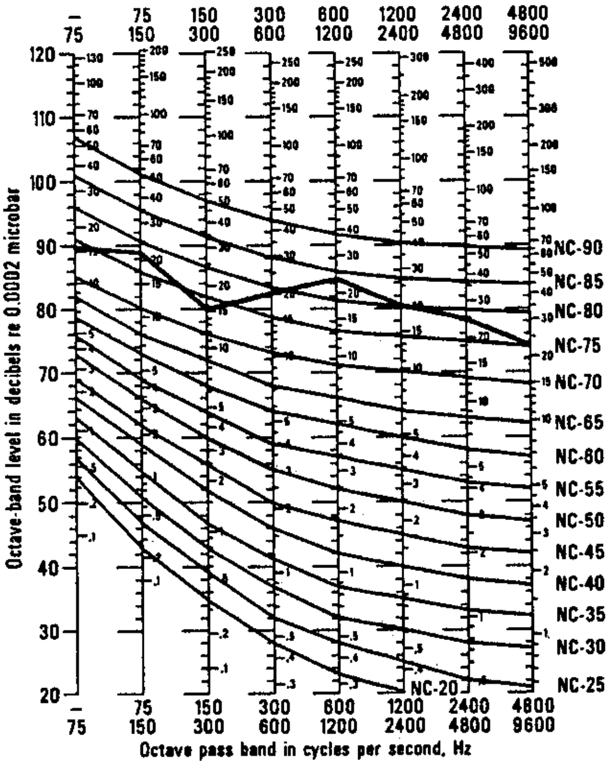
A Scale for levels at about 40 dB
B Scale – 55-85 dB
C Scale – above 85 dB

Sound Pressure (microbars)	Sound Pressure Level (db re 0.002 microbar)	Example
0.0002	0	Threshold of hearing
0.00063	10	
0.002	20	
0.0063	30	
0.02	40	
0.063	50	Residence
0.2	60	Conversational speech
0.63	70	
1.0	74	
2.0	80	
6.3	90	Subway
20	100	Looms in textile mill
63	110	Woodworking
200	120	Hydraulic press
2000	140	Jet plane

Band No.	Center Frequency (Hz)	Frequency Range (Hz)
1	63	37.5 - 75.0
2	125	75.0 - 150
3	250	150 - 300
4	500	300 - 600
5	1000	600 - 1200
6	2000	1200 - 2400
7	4000	2400 - 4800
8	8000	4800 - 9600

Use of these scales is not satisfactory for total analysis of a noise problem. Two different types of sounds in the same scale rating can possess characteristics that have a different effect on the human ear.

A set of contours has been developed known as NC curves (noise criteria curves). A noise criteria curve example follows, accompanied by an octave band analysis of a compressor room. The heavy line shows the latter analysis superimposed on the NC curves. The highest NC value determines the level of concern.



Noise criterion "NC" curves

Octave Band (Hz)	Sound Pressure Level (SPL) (dB re 0.0002 microbar)
31.5-75	90
75-150	89
150-300	80
300-600	82
600-1200	85
1200-2400	80
2400-4800	78
4800-9600	74

Shown below are noise criteria levels in various environments that have proven acceptable.

Space	Noise Criteria	
	Low	High
General office areas	35	50
Executive offices	30	40
Conference rooms	25	35
Secretarial offices	25	35
Accounting areas	35	50
Large engineering and drafting offices	35	50
Assembly or meeting rooms	30	35
Laboratories	35	45
Cafeterias	40	50
Guard houses	35	45
First aid rooms	30	40
Control rooms	35	45
Foremen's rooms and offices - manufacturing	40	50
Corridors and hallways	35	45
Kitchens	40	50
Locker rooms	35	45
Computer rooms	40	60
Private offices	30	45
Board rooms	20	30
Office machinery rooms	40	60
Stenographic rooms	35	50
Laundries	40	60
Supervisor room	30	45
Lobbies	35	45
Mechanical equipment rooms	55	65
Switchgear rooms	50	60
Tank rooms (metallurgical)	50	70
Blower and compressor rooms	60	70
Tool maintenance	45	70
General storage	50	65
Light machinery assembly	55	70
Machine shops	45	65
Plating shops	60	70
Foundries and heavy manufacturing	55	75
Rock crushing and grinding	50	75
Punch press shops	60	75
Chipping and grinding areas	60	75
Shot blast rooms	65	75
Electric furnace rooms	60	75
Pulverizing ball mill rooms	65	75
Riveting operations	65	75
Heat treatment rooms	60	75
Boiler rooms	65	75
Die rooms	65	75
Inspection rooms	55	65
Shipping rooms	50	65
Burning-off area cutting torches	60	70
Lunch room - crusher building	50	60
Jaw crusher building - one crusher	65	75
Lunch area - ball mill floor	50	60
Wash room - ball mill floor	55	65
Diesel engine room	60	70
Electric power station	60	70
High speed blowers - inside	65	75
Crusher building ambient	60	75

The International Standard Organization (ISO) and others have adopted a dBA rating based on frequency and pressure that hopefully simulates the human ear. Several criteria are used. One involves an octave band analysis. The other adjusts pressure for frequency. Shown below are permissible noise exposures based on this latter criteria.

Duration/day in hours	Sound Level dBA Slow Response
8	90
6	92
4	95
3	97
2	100
1.5	102
1	105
0.5	110
0.25 or less	115

The sound abatement procedure must depend on the frequency of the sound. This involves analysis of each machinery item.

Equipment Noise Summary

The noise level of different equipment depends on too many factors for this brief discussion. Within a given type, model may affect noise level. Noise is also dependent on certain operating parameters. The following are typical. They should increase in dB level with the change in the parameter shown.

Internal Combustion Engines: 10 log (hp increase)
30 log (speed increase)

Fans: 10 log (hp increase)
10 log (pressure head increase)

Pumps: 17 log (hp increase)
40 log (speed increase)

Gas Flow: $(\text{velocity})^6$ and $(\text{pressure})^3$

Liquid Flow: 60 log velocity (without cavitation)
120 log velocity (with cavitation)

We must be concerned not only with sound energy transmitted through the air but that transmitted long distances through floors, walls and piping. One needs a "dam" to stop energy flow through solids. It is created by interrupting sound flow with materials possessing high acoustical impedance. Soft rubber and cork have very low impedances. Wood (and water) are rather good. Concrete has an impedance about one-fourth that of steel; glass is about one-third that of steel. Of the common metals cast iron is best.

Centrifugal and axial compressors are major sources of noise. A primary source of noise is the rotor passing by the stators. The frequency of noise depends on rpm, the number of rotors and the number of stationary (diffuser) vanes. There is also the noise source in the inlet and outlet sections. The noise is of a high intensity, and in a frequency range, that propogates readily in connecting piping. It is for this reason that it is almost impossible to make a significant reduction in centrifugal compressor noise unless the piping is treated acoustically.

Inside large turbomachines a wide range of frequencies is generated, usually from 100-6000 hertz. Turbomachinery noise varies widely. In the 4000-12 000 rpm sizes, the dBA level will vary from about 80 for 1000 hp units to about 130 for 30 000 hp units, at a distance of about one meter from the units.

Acoustical lagging is not a panacea to the problem. In some cases a trivial reduction of only 2-3 dBA has resulted. Results vary but they seldom measure up to theoretical performance.

A major source of noise in helical rotor compressors is from the discharge and inlet, in that order. In the discharge proper the level may be 140-145 dB, reduced to about 130 dB, one meter from a unit without a silencer or enclosure.

Increase in inlet and discharge size does not solve the problem in turbomachinery. A large opening may actually increase noise if shock and turbulence are involved. There are two primary criteria – gas should enter near the center line on its way to the impeller and the velocity should be as low as feasible.

Merely lagging noisy machinery is semi-futile. There may be few dB reduction of the higher frequencies but almost none in the lower frequencies. A good enclosure can attenuate (reduce) sound up to 50 dB or more. A good enclosure must have sufficient mass to absorb the sound force necessary to make it oscillate or vibrate. This mass also requires greater sound kinetic energy to move it.

The one meter distance is a rather standard criterion. As a general rule it is advisable to hold the decibel level to 80 in the octave bands from 75-2400 hertz. From all noise sources the dBA level should be in the mideighties. The table at the top of page 301 shows some of the noise limits imposed in a given case.

Equipment	Octave band center frequency, hz								
	31	63	125	250	500	1000	2000	4000	8000
Furnaces, air-cooled heat exchangers, gas valves compressors, and piping systems	100	96	92	88	85	82	80	80	80
Motors	100	96	92	86	83	83	83	83	83
Liquid valves, pumps, turbines, gears, vents and flares	100	96	92	88	85	85	85	85	85

(Octave band sound pressure levels, decibels re 0.0002 microbar)

Based on this, some specifications have included items like the following:

Aerial coolers – 6 or more blades per fan; limit tip speed to about 3000 m/min; use heavy-duty, well damped shrouding and use tip and hub seals.

Fired heaters – use acoustically lined intake air plenum including camping compound, two 90° turns on inlet air piping, and at least 10 cm of 100 kg/m^3 fiber glass with a screen.

Control valves – use quiet design or in-line silencers when needed.

Control valves are the source of much of the noise attributed to piping systems. Once created this noise travels for a long distance and is increased by a maze of piping. Most noise just downstream from the valve is in the 2000-4000 hertz octave band. With normal noise limits, pressure drop ratios less than 2:1 are usually recommended. This may involve using two valves in series.

The shear forces developed in turbulent flow are the primary source of noise in control valves. This aerodynamic noise should be limited to about 90 dBA over an ordinary eight-hour work period. New methods are now available to predict this noise level within ±5%.(9.22) The calculation incorporates the following variables: basic turbulent flow sound level, correction for flow coefficient, correction for valve style and pressure-drop ratio, correction for pipe diameter and wall thickness and a correction for outlet pressure.

Sound control is rapidly developing from an art into a science but many vendors are over-optimistic about the effectiveness of certain sound prevention techniques proposed. Review their recommendations carefully.

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APPENDIX 9A

Some Standard Control Symbols

All lines shall be fine in relation to process piping lines.

(1) Connection to process, or mechanical link, or instrument supply ————————

(2) Pneumatic signal , or undefined signal for process flow diagrams —//——//——//—

(3) Electric signal - - - - - - - -

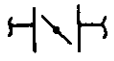
(4) Capillary tubing (filled system) ✱———✱———✱

(5) Hydraulic signal ┕———┕———┕

(6) Electromagnetic or sonic signal (without wiring or tubing) ∿∿∿

AS	Air Supply
ES	Electric Supply
GS	Gas Supply
HS	Hydraulic Supply
NS	Nitrogen Supply
SS	Steam Supply
WS	Water Supply

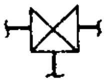
GLOBE, GATE, OR OTHER IN-LINE TYPE NOT OTHERWISE IDENTIFIED



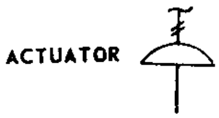
BUTTERFLY, DAMPER, OR LOUVER

ROTARY PLUG OR BALL

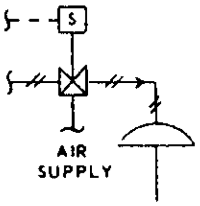
ANGLE

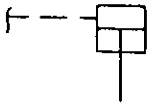


THREE-WAY

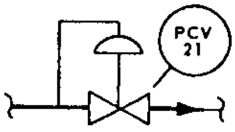


WITHOUT POSITIONER OR OTHER PILOT

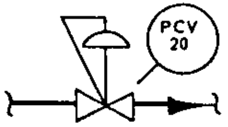




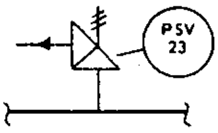
PREFERRED FOR ANY CYLINDER THAT IS ASSEMBLED WITH PILOT SO THAT ASSEMBLY IS ACTUATED BY ONE CONTROLLED INPUT



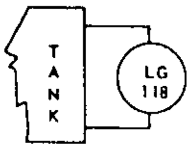
BACKPRESSURE REGULATOR WITH EXTERNAL PRESSURE TAP



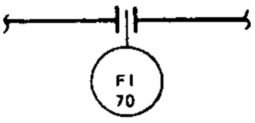
BACKPRESSURE REGULATOR, SELF-CONTAINED



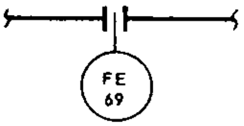
PRESSURE RELIEF OR SAFETY VALVE, ANGLE PATTERN, SPRING- OR WEIGHT-LOADED, OR WITH INTEGRAL PILOT



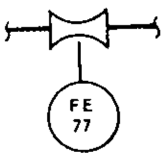
GAGE GLASS, EXTERNALLY CONNECTED



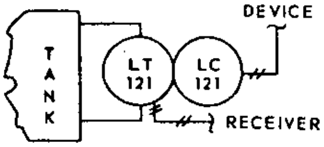
ORIFICE PLATE WITH FLANGE OR CORNER TAPS CONNECTED TO DIFFERENTIAL-PRESSURE-TYPE FLOW INDICATOR



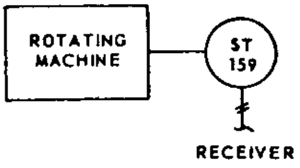
ORIFICE PLATE WITH FLANGE OR CORNER TAPS



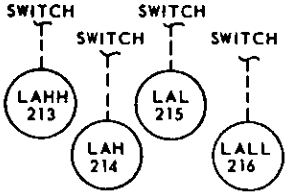
VENTURI TUBE OR FLOW NOZZLE



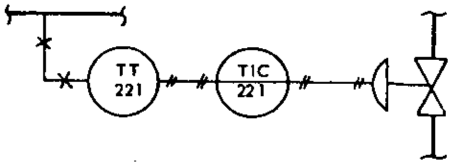
DUPLEX LEVEL TRANSMITTER-CONTROLLER, EXTERNAL-FLOAT OR EXTERNAL-DISPLACER TYPE



SPEED TRANSMITTER



SEPARATE ALARMS FOR ABNORMAL LEVEL – VERY HIGH, HIGH, LOW, AND VERY LOW



TEMPERATURE INDICATING CONTROLLER THAT IS EITHER SELF-CONTAINED OR A CONTROL STATION PLUG-IN CONTROLLER COMBINATION

FLANGE TAPS—BASIC ORIFICE FACTORS—F_b

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	2			3				4	
	1.689	1.939	2.067	2.300	2.626	2.900	3.068	3.152	3.438
.250	12.695	12.708	12.711	12.714	12.712	12.708	12.705	12.703	12.697
.375	28.474	28.440	28.427	28.411	28.393	28.382	28.376	28.373	28.364
.500	50.777	50.587	50.521	50.435	50.356	50.313	50.292	50.283	50.258
.625	80.090	79.508	79.311	79.052	78.817	78.687	78.625	78.599	78.523
.750	117.09	115.62	115.14	114.52	113.99	113.70	113.56	113.50	113.33
.875	162.95	159.56	158.47	157.12	156.00	155.41	155.14	155.03	154.71
1.000	219.77	212.47	210.22	207.44	205.18	204.04	203.54	203.33	202.75
1.125	290.99	276.19	271.70	266.35	262.06	259.95	259.04	258.65	257.63
1.250	385.78	353.58	345.13	335.12	327.39	323.63	322.03	321.37	319.61
1.375	....	448.59	433.50	415.75	402.18	395.80	393.09	391.97	389.03
1.500	....	....	542.27	510.86	487.98	477.36	472.96	471.14	466.39
1.625	....	....	....	623.91	586.82	569.65	562.58	559.72	552.31
1.750	....	....	....	....	701.26	674.44	663.42	658.96	647.54
1.875	....	....	....	....	834.88	793.88	777.18	770.44	753.17
2.000	....	....	....	....	....	930.65	906.01	896.06	870.59
2.125	....	....	....	....	....	1091.2	1052.5	1038.1	1001.4
2.250	....	....	....	....	....	....	1223.2	1199.9	1147.7
2.375	....	....	....	....	....	....	....	....	1311.7
2.500	....	....	....	....	....	....	....	....	1498.4

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	4		6				8		
	3.826	4.026	4.897	5.189	5.761	6.065	7.625	7.981	8.071
.250	12.688	12.683	....	....	....	....	....	....	....
.375	28.353	28.348	....	....	....	....	....	....	....
.500	50.234	50.224	50.197	50.191	50.182	50.178	....	....	....
.625	78.450	78.421	78.338	78.321	78.296	78.287	....	....	....
.750	113.15	113.08	112.87	112.82	112.75	112.72	....	....	....
.875	154.40	154.27	153.88	153.78	153.63	153.56	153.34	153.31	153.31
1.000	202.20	201.99	201.34	201.19	200.96	200.85	200.46	200.40	200.38
1.125	256.69	256.33	255.31	255.08	254.72	254.56	253.99	253.89	253.87
1.250	318.03	317.45	315.83	315.48	314.95	314.72	313.91	313.78	313.74
1.375	386.45	385.51	382.99	382.47	381.70	381.37	380.25	380.06	380.02
1.500	462.27	460.79	456.93	456.16	455.03	454.57	453.02	452.78	452.72
1.625	545.89	543.61	537.77	536.64	535.03	534.38	532.27	531.95	531.87
1.750	637.83	634.39	625.73	624.09	621.79	620.89	618.02	617.60	617.50
1.875	738.75	733.68	721.03	718.69	715.44	714.18	710.32	709.77	709.64
2.000	849.41	842.12	823.99	820.68	816.13	814.41	809.22	808.50	808.34
2.125	970.95	960.48	934.97	930.35	924.07	921.71	914.78	913.85	913.64
2.250	1104.7	1089.9	1054.4	1048.1	1039.5	1036.3	1027.1	1025.9	1025.6
2.375	1252.1	1231.7	1182.9	1174.2	1162.6	1158.3	1146.2	1144.7	1144.3
2.500	1415.0	1387.2	1320.9	1309.3	1293.8	1288.2	1272.3	1270.3	1269.8
2.625	1595.6	1558.2	1469.2	1453.9	1433.5	1426.0	1405.4	1402.9	1402.3
2.750	1797.1	1746.7	1628.9	1608.7	1582.0	1572.3	1545.7	1542.5	1541.8
2.875	....	1955.5	1801.0	1774.5	1740.0	1727.5	1693.4	1689.3	1688.4
3.000	....	2195.0	1986.6	1952.4	1907.8	1891.9	1848.6	1843.5	1842.3
3.125	....	....	2187.2	2143.4	2086.4	2066.1	2011.6	2005.2	2003.8
3.250	....	....	2404.2	2348.8	2276.5	2250.8	2182.6	2174.6	2172.9
3.375	....	....	2639.5	2569.8	2479.1	2446.8	2361.8	2352.0	2349.9
3.500	....	....	2895.5	2808.1	2695.1	2654.9	2549.7	2537.7	2535.0
3.625	....	....	3180.8	3065.3	2925.7	2876.0	2746.5	2731.8	2728.6
3.750	....	....	....	3345.5	3172.1	3111.2	2952.6	2934.8	2930.8
3.875	....	....	....	3657.7	3435.8	3361.5	3168.3	3146.9	3142.1
4.000	....	....	....	....	3718.2	3628.2	3394.3	3368.5	3362.9
4.250	....	....	....	....	4354.8	4216.6	3879.4	3842.3	3834.2
4.500	....	....	....	....	....	4900.9	4412.8	4360.5	4349.0
4.750	....	....	....	....	....	....	5000.7	4928.1	4912.2
5.000	....	....	....	....	....	....	5650.0	5551.1	5529.5
5.250	....	....	....	....	....	....	6369.3	6236.4	6207.3
5.500	....	....	....	....	....	....	7170.9	6992.0	6953.6
5.750	....	....	....	....	....	....	....	7830.0	7777.8
6.000	....	....	....	....	....	....	....	....	8707.0

APPENDIX 9B

FLANGE TAPS—BASIC ORIFICE FACTORS—F_b

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	10			12			16		
	9.564	10.020	10.136	11.376	11.938	12.090	14.688	15.000	15.250
1.000	200.20	....	....	....	....	....	....	....	....
1.125	253.56	253.48	253.47	....	....	....	....	....	....
1.250	313.31	313.20	313.18	312.94	312.85	312.83	....	....	....
1.375	379.44	379.29	379.26	378.94	378.82	378.79	....	....	....
1.500	451.95	451.76	451.72	451.30	451.14	451.10	450.53	450.48	....
1.625	530.87	530.63	530.57	530.04	529.84	529.78	529.06	528.99	528.94
1.750	616.20	615.90	615.83	615.16	614.91	614.84	613.94	613.85	613.78
1.875	707.98	707.60	707.51	706.68	706.36	706.28	705.18	705.07	704.99
2.000	806.23	805.76	805.65	804.61	804.23	804.13	802.78	802.65	802.56
2.125	910.97	910.38	910.24	908.98	908.51	908.39	906.77	906.61	906.49
2.250	1022.2	1021.5	1021.3	1019.8	1019.2	1019.1	1017.1	1017.0	1016.8
2.375	1140.1	1139.2	1139.0	1137.1	1136.4	1136.2	1133.9	1133.7	1133.5
2.500	1264.5	1263.4	1263.1	1260.8	1260.0	1259.8	1257.1	1256.8	1256.6
2.625	1395.6	1394.2	1393.9	1391.1	1390.1	1389.9	1386.7	1386.4	1386.1
2.750	1533.4	1531.7	1531.3	1528.0	1526.8	1526.5	1522.7	1522.4	1522.1
2.875	1678.0	1675.9	1675.4	1671.4	1670.0	1669.6	1665.2	1664.8	1664.5
3.000	1829.4	1826.9	1826.3	1821.4	1819.7	1819.3	1814.1	1813.7	1813.3
3.125	1987.8	1984.7	1984.0	1978.1	1976.1	1975.6	1969.6	1969.0	1968.6
3.250	2153.2	2149.5	2148.6	2141.5	2139.2	2138.6	2131.5	2130.9	2130.4
3.375	2325.7	2321.2	2320.2	2311.7	2308.9	2308.2	2299.9	2299.2	2298.7
3.500	2505.6	2500.1	2498.9	2488.7	2485.4	2484.6	2474.9	2474.1	2473.5
3.625	2692.8	2686.2	2684.7	2672.6	2668.7	2667.7	2656.4	2655.5	2654.8
3.750	2887.6	2879.7	2877.9	2863.5	2858.8	2857.7	2844.6	2843.5	2842.7
3.875	3090.1	3080.7	3078.5	3061.4	3055.9	3054.6	3039.4	3038.1	3037.2
4.000	3300.6	3289.3	3286.8	3266.4	3260.0	3258.5	3240.8	3239.4	3238.3
4.250	3746.1	3730.2	3726.7	3698.4	3689.6	3687.5	3663.8	3661.9	3660.5
4.500	4226.0	4204.1	4199.2	4160.4	4148.4	4145.5	4113.9	4111.5	4109.7
4.750	4742.7	4712.8	4706.2	4653.4	4637.2	4633.4	4591.5	4588.4	4586.0
5.000	5298.6	5258.4	5249.6	5179.0	5157.4	5152.3	5097.2	5093.1	5090.1
5.250	5897.4	5843.6	5831.8	5738.5	5710.0	5703.3	5631.4	5626.1	5622.2
5.500	6543.1	6471.9	6456.3	6333.8	6296.6	6287.9	6194.8	6188.1	6183.1
5.750	7240.0	7146.9	7126.5	6966.9	6919.0	6907.8	6788.1	6779.6	6773.3
6.000	7993.3	7872.9	7846.6	7640.4	7579.0	7564.7	7412.3	7401.5	7393.6
6.250	8808.9	8654.8	8621.1	8357.3	8278.9	8260.7	8068.3	8054.8	8044.8
6.500	9693.3	9498.1	9455.3	9121.0	9021.7	8998.7	8757.3	8740.3	8727.9
6.750	10654	10409	10355	9935.2	9810.5	9781.6	9480.4	9459.4	9444.0
7.000	11711	11394	11327	10804	10649	10613	10239	10213	10194
7.250	....	12467	12381	11732	11540	11496	11035	11003	10980
7.500	....	13656	13541	12725	12489	12434	11869	11831	11803
7.750	....	....	....	13787	13500	13433	12745	12698	12664
8.000	....	....	....	14927	14578	14498	13664	13607	13566
8.250	....	....	....	16158	15730	15633	14628	14560	14510
8.500	....	....	....	17505	16963	16845	15642	15560	15500
8.750	....	....	....	....	18297	18148	16706	16609	16538
9.000	....	....	....	....	....	19566	17826	17711	17627
9.250	....	....	....	....	....	....	19004	18868	18769
9.500	....	....	....	....	....	....	20245	20085	19969
9.750	....	....	....	....	....	....	21552	21365	21229
10.000	....	....	....	....	....	....	22930	22712	22554
10.250	....	....	....	....	....	....	24385	24132	23947
10.500	....	....	....	....	....	....	25924	25628	25414
10.750	....	....	....	....	....	....	27567	27210	26960
11.000	....	....	....	....	....	....	29331	28899	28598
11.250	....	....	....	....	....	....	....	30710	30346

FLANGE TAPS—BASIC ORIFICE FACTORS—F_b

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	20			24			30		
	18.814	19.000	19.250	22.626	23.000	23.250	28.628	29.000	29.250
2.000	801.40	801.35	801.29	....	....	....	....	....	....
2.125	905.11	905.06	904.98	....	....	....	....	....	....
2.250	1015.2	1015.1	1015.0	....	....	....	....	....	....
2.375	1131.6	1131.5	1131.4	1130.2	1130.1	1130.0	....	....	....
2.500	1254.4	1254.3	1254.2	1252.8	1252.6	1252.6	....	....	....
2.625	1383.6	1383.5	1383.3	1381.7	1381.5	1381.4	....	....	....
2.750	1519.1	1519.0	1518.8	1517.0	1516.8	1516.7	....	....	....
2.875	1661.0	1660.9	1660.7	1658.6	1658.4	1658.3	1656.0	....	....
3.000	1809.4	1809.2	1809.0	1806.6	1806.4	1806.2	1803.7	1803.5	1803.4
3.125	1964.1	1963.9	1963.7	1961.0	1960.7	1960.6	1957.7	1957.5	1957.4
3.250	2125.3	2125.1	2124.8	2121.7	2121.5	2121.3	2118.0	2117.9	2117.7
3.375	2292.9	2292.6	2292.3	2288.9	2288.6	2288.4	2284.7	2284.5	2284.4
3.500	2466.9	2466.6	2466.3	2462.4	2462.1	2461.8	2457.8	2457.6	2457.4
3.625	2647.3	2647.0	2646.6	2642.4	2642.0	2641.7	2637.3	2637.0	2636.8
3.750	2834.2	2833.9	2833.5	2828.7	2828.3	2828.0	2823.1	2822.8	2822.6
3.875	3027.5	3027.3	3026.8	3021.5	3021.0	3020.7	3015.2	3014.9	3014.7
4.000	3227.5	3227.1	3226.5	3220.6	3220.1	3219.8	3213.8	3213.5	3213.3
4.250	3646.7	3646.2	3645.6	3638.3	3637.7	3637.2	3630.1	3629.7	3629.4
4.500	4092.1	4091.5	4090.6	4081.8	4081.0	4080.5	4071.9	4071.4	4071.1
4.750	4563.7	4562.9	4561.9	4551.1	4550.1	4549.5	4539.3	4538.8	4538.4
5.000	5061.8	5060.8	5059.6	5046.4	5045.2	5044.5	5032.4	5031.8	5031.4
5.250	5586.6	5585.4	5583.8	5567.7	5566.4	5565.5	5551.3	5550.5	5550.0
5.500	6138.2	6136.7	6134.8	6115.3	6113.6	6112.6	6095.8	6094.9	6094.4
5.750	6717.1	6715.2	6712.8	6689.1	6687.2	6685.9	6666.2	6665.2	6664.5
6.000	7323.4	7321.1	7318.2	7289.4	7287.1	7285.6	7262.5	7261.3	7260.5
6.250	7957.5	7954.7	7951.2	7916.4	7913.7	7911.9	7884.7	7883.4	7882.5
6.500	8619.9	8616.5	8612.2	8570.2	8566.9	8564.8	8533.0	8531.4	8530.4
6.750	9311.1	9306.9	9301.6	9251.1	9247.2	9244.7	9207.4	9205.6	9204.4
7.000	10031	10026	10020	9959.3	9954.6	9951.7	9908.0	9905.9	9904.6
7.250	10782	10776	10768	10695	10689	10686	10635	10633	10631
7.500	11562	11555	11546	11459	11452	11448	11388	11386	11384
7.750	12374	12365	12354	12250	12243	12238	12168	12165	12163
8.000	13218	13207	13194	13071	13062	13056	12975	12971	12969
8.250	14095	14082	14066	13920	13910	13903	13809	13805	13802
8.500	15005	14990	14971	14799	14787	14779	14669	14665	14661
8.750	15950	15933	15911	15708	15693	15684	15557	15552	15548
9.000	16932	16911	16885	16647	16630	16620	16473	16466	16462
9.250	17950	17926	17895	17618	17598	17585	17416	17409	17404
9.500	19007	18979	18943	18620	18597	18582	18387	18379	18373
9.750	20104	20071	20030	19655	19628	19611	19386	19377	19371
10.000	21243	21205	21157	20723	20692	20672	20414	20403	20396
10.250	22426	22382	23325	21825	21789	21767	21471	21458	21450
10.500	23654	23603	23538	22962	22921	22895	22556	22542	22533
10.750	24931	24872	24797	24134	24088	24058	23672	23656	23646
11.000	26257	26190	26104	25344	25290	25257	24817	24799	24787
11.250	27636	27559	27460	26592	26531	26492	25992	25972	25959
11.500	29070	28982	28870	27878	27809	27766	27199	27176	27161
11.750	30563	30462	30334	29205	29126	29077	28436	28411	28394
12.000	32116	32001	31856	30574	30485	30429	29706	29677	29659
12.500	35417	35270	35084	33444	33330	33259	32343	32306	32283
13.000	39003	38817	38581	36502	36357	36267	35114	35068	35039
13.500	42913	42673	42375	39762	39581	39467	38025	37968	37932
14.000	47244	46921	46523	43241	43015	42874	41082	41012	40968
14.500	....	....	....	46958	46679	46505	44291	44206	44152
15.000	....	....	....	50934	50591	50378	47661	47557	47490
15.500	....	....	....	55192	54774	54513	51202	51075	50994
16.000	....	....	....	59759	59251	58935	54923	54769	54671
16.500	....	....	....	64701	64060	63671	58835	58649	58531
17.000	....	....	....	....	69288	68792	62960	62728	62586
17.500	....	....	....	....	....	....	67282	67017	66848
18.000	....	....	....	....	....	....	71844	71530	71330
18.500	....	....	....	....	....	....	76653	76282	76046
19.000	....	....	....	....	....	....	81725	81289	81011
19.500	....	....	....	....	....	....	87078	86568	86244
20.000	....	....	....	....	....	....	92734	92140	91761
20.500	....	....	....	....	....	....	98728	98025	97584
21.000	....	....	....	....	....	....	105134	104282	103752
21.500	....	....	....	....	....	....	....	110983	110340

"b" VALUES FOR REYNOLDS NUMBER FACTOR, F_r,—FLANGE TAPS

$$F_r = 1 + \frac{b}{\sqrt{h_w p_f}}$$

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	2			3				4	
	1.689	1.939	2.067	2.300	2.626	2.900	3.068	3.152	3.438
.250	.0879	.0911	.0926	.0960	.0979	.0999	.1010	.1014	.1030
.375	.0677	.0709	.0726	.0755	.0792	.0820	.0836	.0844	.0867
.500	.0562	.0576	.0588	.0612	.0648	.0677	.0696	.0703	.0730
.625	.0520	.0505	.0506	.0516	.0541	.0566	.0583	.0591	.0618
.750	.0536	.0485	.0471	.0482	.0470	.0486	.0498	.0504	.0528
.875	.0595	.0506	.0478	.0445	.0429	.0433	.0438	.0442	.0460
1.000	.0677	.0559	.0515	.0458	.0416	.0403	.0402	.0403	.0411
1.125	.0762	.0630	.0674	.0495	.0427	.0398	.0386	.0383	.0380
1.250	.0824	.0707	.0646	.0550	.0456	.0408	.0388	.0381	.0366
1.375	....	.0772	.0715	.0614	.0501	.0435	.0406	.0394	.0365
1.500	....	....	.0773	.0679	.0554	.0474	.0436	.0420	.0378
1.625	....	....	....	.0735	.0613	.0522	.0477	.0457	.0402
1.750	....	....	....	....	.0669	.0575	.0524	.0500	.0434
1.875	....	....	....	....	.0717	.0628	.0574	.0549	.0473
2.000	....	....	....	....	....	.0676	.0624	.0598	.0517
2.125	....	....	....	....	....	.0715	.0669	.0644	.0563
2.250	....	....	....	....	....	....	.0706	.0686	.0607
2.375	....	....	....	....	....	....	....	....	.0648
2.500	....	....	....	....	....	....	....	....	.0683

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	4		6				8		
	3.826	4.026	4.897	5.189	5.761	6.065	7.625	7.981	8.071
.250	.1047	.1054	....	....	....	....			
.375	.0894	.0907	....	....	....	....			
.500	.0763	.0779	.0836	.0852	.0880	.0892	....	....	....
.625	.0653	.0670	.0734	.0753	.0786	.0801	....	....	....
.750	.0561	.0578	.0645	.0665	.0701	.0718	....	....	....
.875	.0487	.0502	.0567	.0587	.0625	.0643	.0723	.0738	.0742
1.000	.0430	.0442	.0500	.0520	.0557	.0576	.0660	.0676	.0680
1.125	.0388	.0398	.0444	.0462	.0498	.0517	.0602	.0619	.0623
1.250	.0361	.0364	.0399	.0414	.0447	.0464	.0549	.0566	.0571
1.375	.0347	.0344	.0363	.0375	.0403	.0419	.0501	.0518	.0523
1.500	.0346	.0336	.0336	.0344	.0367	.0381	.0457	.0474	.0479
1.625	.0354	.0338	.0318	.0322	.0337	.0348	.0418	.0435	.0439
1.750	.0372	.0350	.0307	.0306	.0314	.0322	.0383	.0399	.0403
1.875	.0398	.0370	.0305	.0298	.0298	.0303	.0353	.0366	.0371
2.000	.0430	.0396	.0308	.0296	.0287	.0288	.0327	.0340	.0343
2.125	.0467	.0427	.0318	.0300	.0281	.0278	.0304	.0315	.0318
2.250	.0507	.0462	.0334	.0310	.0281	.0274	.0286	.0296	.0297
2.375	.0548	.0501	.0354	.0324	.0286	.0274	.0271	.0278	.0280
2.500	.0589	.0540	.0378	.0342	.0296	.0279	.0259	.0264	.0265
2.625	.0628	.0579	.0406	.0365	.0308	.0287	.0251	.0253	.0254
2.750	.0669	.0615	.0436	.0391	.0324	.0300	.0248	.0246	.0246
2.875	....	.0647	.0468	.0418	.0343	.0314	.0244	.0240	.0240
3.000	....	.0673	.0500	.0448	.0366	.0332	.0246	.0238	.0237
3.125	....	....	.0533	.0479	.0389	.0353	.0248	.0239	.0237
3.250	....	....	.0564	.0510	.0416	.0375	.0254	.0242	.0240
3.375	....	....	.0594	.0541	.0443	.0400	.0263	.0248	.0244
3.500	....	....	.0620	.0569	.0472	.0426	.0273	.0256	.0251
3.625	....	....	.0643	.0597	.0500	.0452	.0286	.0265	.0260
3.750	....	....	....	.0621	.0527	.0479	.0300	.0276	.0271
3.875	....	....	....	.0640	.0553	.0506	.0316	.0289	.0283
4.000	....	....	....	....	.0578	.0531	.0334	.0304	.0297
4.250	....	....	....	....	.0620	.0579	.0372	.0338	.0330
4.500	....	....	....	....	....	.0618	.0414	.0375	.0368
4.750	....	....	....	....	....	....	.0457	.0416	.0405
5.000	....	....	....	....	....	....	.0500	.0457	.0446
5.250	....	....	....	....	....	....	.0539	.0497	.0487
5.500	....	....	....	....	....	....	.0574	.0535	.0524
5.750	....	....	....	....	....	....	....	.0569	.0559
6.000	....	....	....	....	....	....	....	....	.0594

“b” VALUES FOR
REYNOLDS NUMBER FACTOR, F_r,—FLANGE TAPS

$$F_r = 1 + \frac{b}{\sqrt{h_w\, p_f}}$$

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	10			12			16		
	9.564	10.020	10.136	11.376	11.938	12.090	14.688	15.000	15.250
1.000	.0738	....	....	....	....	....	....	....	....
1.125	.0685	.0701	.0705	....	....	....	....	....	....
1.250	.0636	.0652	.0656	.0698	.0714	.0718	....	....	....
1.375	.0588	.0606	.0610	.0654	.0671	.0676	....	....	....
1.500	.0545	.0563	.0568	.0612	.0631	.0636	.0706	.0713	....
1.625	.0504	.0523	.0527	.0573	.0592	.0597	.0670	.0678	.0684
1.750	.0467	.0486	.0490	.0536	.0556	.0560	.0636	.0644	.0650
1.875	.0433	.0451	.0455	.0501	.0521	.0526	.0604	.0612	.0618
2.000	.0401	.0419	.0423	.0469	.0488	.0492	.0572	.0581	.0587
2.125	.0372	.0389	.0393	.0438	.0458	.0463	.0542	.0551	.0558
2.250	.0346	.0362	.0366	.0410	.0429	.0434	.0514	.0523	.0529
2.375	.0322	.0337	.0341	.0383	.0402	.0407	.0487	.0496	.0502
2.500	.0302	.0315	.0319	.0359	.0377	.0382	.0461	.0470	.0476
2.625	.0283	.0296	.0299	.0336	.0354	.0358	.0436	.0445	.0452
2.750	.0267	.0278	.0281	.0316	.0332	.0336	.0413	.0422	.0428
2.875	.0254	.0263	.0265	.0297	.0312	.0317	.0391	.0399	.0406
3.000	.0243	.0250	.0252	.0278	.0294	.0298	.0370	.0378	.0385
3.125	.0234	.0239	.0241	.0264	.0278	.0282	.0350	.0358	.0365
3.250	.0226	.0230	.0231	.0251	.0263	.0266	.0331	.0339	.0346
3.375	.0221	.0223	.0224	.0239	.0250	.0253	.0314	.0321	.0328
3.500	.0219	.0218	.0218	.0229	.0238	.0241	.0298	.0305	.0311
3.625	.0218	.0214	.0214	.0221	.0228	.0230	.0282	.0290	.0296
3.750	.0218	.0213	.0212	.0214	.0219	.0221	.0268	.0275	.0281
3.875	.0221	.0213	.0211	.0208	.0212	.0213	.0255	.0262	.0267
4.000	.0225	.0214	.0212	.0204	.0206	.0207	.0243	.0249	.0254
4.250	.0238	.0222	.0219	.0200	.0198	.0196	.0223	.0228	.0232
4.500	.0256	.0236	.0231	.0201	.0195	.0194	.0206	.0210	.0213
4.750	.0279	.0254	.0248	.0207	.0196	.0194	.0193	.0196	.0198
5.000	.0307	.0277	.0270	.0217	.0202	.0198	.0184	.0186	.0187
5.250	.0337	.0303	.0296	.0231	.0212	.0208	.0178	.0178	.0179
5.500	.0370	.0332	.0323	.0249	.0226	.0221	.0176	.0174	.0174
5.750	.0404	.0363	.0354	.0270	.0243	.0237	.0176	.0174	.0172
6.000	.0438	.0395	.0386	.0294	.0263	.0256	.0180	.0176	.0173
6.250	.0473	.0429	.0418	.0320	.0285	.0277	.0186	.0180	.0177
6.500	.0505	.0462	.0451	.0347	.0309	.0300	.0195	.0186	.0183
6.750	.0536	.0493	.0483	.0376	.0335	.0325	.0206	.0196	.0192
7.000	.0562	.0523	.0513	.0406	.0362	.0351	.0220	.0210	.0202
7.250	....	.0550	.0540	.0436	.0390	.0379	.0235	.0224	.0216
7.500	....	.0572	.0564	.0463	.0418	.0407	.0252	.0240	.0230
7.750	....	....	....	.0491	.0446	.0434	.0271	.0257	.0246
8.000	....	....	....	.0517	.0473	.0461	.0291	.0276	.0264
8.250	....	....	....	.0540	.0498	.0487	.0312	.0296	.0283
8.500	....	....	....	.0560	.0522	.0511	.0334	.0317	.0303
8.750	....	....	....	....	.0543	.0534	.0357	.0338	.0324
9.000	....	....	....	....	....	.0553	.0380	.0361	.0346
9.250	....	....	....	....	....	....	.0402	.0383	.0368
9.500	....	....	....	....	....	....	.0425	.0406	.0390
9.750	....	....	....	....	....	....	.0447	.0428	.0412
10.000	....	....	....	....	....	....	.0468	.0449	.0434
10.250	....	....	....	....	....	....	.0489	.0470	.0456
10.500	....	....	....	....	....	....	.0508	.0490	.0475
10.750	....	....	....	....	....	....	.0526	.0509	.0495
11.000	....	....	....	....	....	....	.0541	.0526	.0513
11.250	....	....	....	....	....	....	....	.0541	.0528

“b” VALUES FOR
REYNOLDS NUMBER FACTOR, F_r,—FLANGE TAPS

$$F_r = 1 + \frac{b}{\sqrt{h_w\, p_f}}$$

Orifice Diam., d, Inches	Internal Diameter of Pipe, D, Inches								
	20			24			30		
	18.814	19.000	19.250	22.626	23.000	23.250	28.628	29.000	29.250
2.000	.0667	.0671	.0676	....	....	....	....	....	....
2.125	.0640	.0644	.0649	....	....	....	....	....	....
2.250	.0614	.0618	.0622	....	....	....	....	....	....
2.375	.0588	.0592	.0597	.0659	.0665	.0669	....	....	....
2.500	.0563	.0568	.0573	.0636	.0642	.0646	....	....	....
2.625	.0540	.0544	.0549	.0614	.0620	.0624	....	....	....
2.750	.0517	.0521	.0526	.0592	.0599	.0603	....	....	....
2.875	.0494	.0499	.0504	.0571	.0578	.0582	.0662	....	....
3.000	.0473	.0477	.0483	.0551	.0557	.0562	.0644	.0649	.0652
3.125	.0452	.0457	.0462	.0531	.0538	.0542	.0626	.0631	.0634
3.250	.0433	.0437	.0442	.0511	.0518	.0523	.0608	.0613	.0616
3.375	.0414	.0418	.0423	.0493	.0500	.0504	.0590	.0596	.0599
3.500	.0395	.0399	.0405	.0474	.0481	.0486	.0574	.0579	.0582
3.625	.0378	.0382	.0387	.0457	.0464	.0468	.0557	.0562	.0566
3.750	.0361	.0365	.0370	.0440	.0447	.0451	.0541	.0546	.0550
3.875	.0345	.0349	.0354	.0423	.0430	.0435	.0525	.0530	.0534
4.000	.0329	.0333	.0339	.0407	.0414	.0419	.0509	.0515	.0518
4.250	.0301	.0304	.0310	.0376	.0384	.0388	.0479	.0485	.0488
4.500	.0275	.0279	.0283	.0348	.0356	.0360	.0450	.0456	.0460
4.750	.0252	.0256	.0260	.0322	.0328	.0333	.0423	.0429	.0433
5.000	.0232	.0235	.0239	.0297	.0304	.0308	.0397	.0403	.0407
5.250	.0214	.0217	.0220	.0275	.0281	.0285	.0373	.0378	.0382
5.500	.0199	.0201	.0204	.0254	.0260	.0264	.0349	.0355	.0359
5.750	.0186	.0188	.0191	.0236	.0241	.0245	.0327	.0333	.0337
6.000	.0176	.0177	.0179	.0219	.0224	.0228	.0306	.0312	.0316
6.250	.0167	.0168	.0170	.0204	.0208	.0212	.0287	.0292	.0296
6.500	.0161	.0162	.0163	.0191	.0195	.0198	.0269	.0274	.0277
6.750	.0157	.0157	.0157	.0179	.0183	.0185	.0252	.0257	.0260
7.000	.0155	.0155	.0154	.0169	.0172	.0174	.0236	.0240	.0244
7.250	.0155	.0154	.0153	.0161	.0163	.0165	.0221	.0226	.0229
7.500	.0157	.0155	.0154	.0154	.0156	.0157	.0208	.0212	.0215
7.750	.0160	.0158	.0156	.0148	.0150	.0151	.0195	.0199	.0202
8.000	.0166	.0163	.0160	.0144	.0145	.0146	.0184	.0187	.0190
8.250	.0172	.0169	.0165	.0142	.0142	.0142	.0174	.0177	.0179
8.500	.0180	.0177	.0172	.0141	.0140	.0140	.0164	.0168	.0170
8.750	.0190	.0186	.0180	.0141	.0140	.0139	.0156	.0159	.0161
9.000	.0201	.0196	.0190	.0143	.0141	.0140	.0149	.0152	.0153
9.250	.0213	.0208	.0201	.0146	.0143	.0141	.0143	.0145	.0146
9.500	.0226	.0220	.0213	.0150	.0146	.0144	.0138	.0139	.0141
9.750	.0240	.0234	.0226	.0155	.0150	.0147	.0133	.0135	.0136
10.000	.0256	.0249	.0240	.0161	.0155	.0152	.0130	.0131	.0132
10.250	.0271	.0264	.0255	.0168	.0162	.0158	.0128	.0128	.0128
10.500	.0288	.0280	.0270	.0176	.0169	.0164	.0126	.0126	.0126
10.750	.0305	.0297	.0286	.0185	.0176	.0172	.0125	.0125	.0125
11.000	.0322	.0314	.0303	.0194	.0186	.0181	.0125	.0124	.0124
11.250	.0340	.0332	.0320	.0205	.0196	.0190	.0126	.0125	.0124
11.500	.0358	.0349	.0338	.0216	.0207	.0200	.0128	.0126	.0125
11.750	.0376	.0367	.0355	.0228	.0218	.0211	.0130	.0128	.0127
12.000	.0394	.0385	.0373	.0241	.0230	.0223	.0134	.0131	.0129
12.500	.0429	.0420	.0408	.0267	.0255	.0248	.0142	.0138	.0136
13.000	.0463	.0454	.0442	.0296	.0282	.0274	.0153	.0148	.0145
13.500	.0494	.0485	.0474	.0326	.0311	.0302	.0166	.0160	.0157
14.000	.0520	.0512	.0502	.0356	.0341	.0331	.0182	.0175	.0171
14.500	....	....	....	.0386	.0370	.0360	.0199	.0192	.0187
15.000	....	....	....	.0415	.0400	.0390	.0218	.0209	.0204
15.500	....	....	....	.0443	.0428	.0418	.0239	.0230	.0224
16.000	....	....	....	.0470	.0455	.0446	.0260	.0250	.0244
16.500	....	....	....	.0494	.0480	.0471	.0283	.0273	.0266
17.000	....	....	....	....	.0503	.0494	.0307	.0296	.0288
17.500	....	....	....	....	....	....	.0331	.0319	.0312
18.000	....	....	....	....	....	....	.0355	.0343	.0335
18.500	....	....	....	....	....	....	.0379	.0366	.0358
19.000	....	....	....	....	....	....	.0402	.0390	.0382
19.500	....	....	....	....	....	....	.0424	.0412	.0404
20.000	....	....	....	....	....	....	.0446	.0434	.0426
20.500	....	....	....	....	....	....	.0466	.0455	.0448
21.000	....	....	....	....	....	....	.0485	.0475	.0467
21.500	....	....	....	....	....	....	....	.0492	.0485

EXPANSION FACTORS — FLANGE TAPS — Y_1

Static Pressure Taken from Upstream Taps

$\frac{h_w}{P_{f1}}$ Ratio	$\beta = \frac{d}{D}$ Ratio												
	.1	.2	.3	.4	.45	.50	.52	.54	.56	.58	.60	.61	.62
0.0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
0.1	.9989	.9989	.9989	.9988	.9988	.9988	.9988	.9988	.9988	.9988	.9987	.9987	.9987
0.2	.9977	.9977	.9977	.9977	.9976	.9976	.9976	.9976	.9975	.9975	.9975	.9975	.9974
0.3	.9966	.9966	.9966	.9955	.9965	.9964	.9964	.9963	.9963	.9963	.9962	.9962	.9962
0.4	.9954	.9954	.9954	.9953	.9953	.9952	.9952	.9951	.9951	.9950	.9949	.9949	.9949
0.5	.9943	.9943	.9943	.9942	.9941	.9940	.9940	.9939	.9938	.9938	.9937	.9936	.9936
0.6	.9932	.9932	.9931	.9930	.9929	.9928	.9927	.9927	.9926	.9925	.9924	.9924	.9923
0.7	.9920	.9920	.9920	.9919	.9918	.9916	.9915	.9915	.9914	.9913	.9912	.9911	.9910
0.8	.9909	.9909	.9908	.9907	.9906	.9904	.9903	.9902	.9901	.9900	.9899	.9898	.9897
0.9	.9898	.9897	.9897	.9895	.9894	.9892	.9891	.9890	.9889	.9888	.9886	.9885	.9885
1.0	.9986	.9886	.9885	.9884	.9882	.9880	.9879	.9878	.9877	.9875	.9874	.9873	.9872
1.1	.9875	.9875	.9874	.9872	.9870	.9868	.9867	.9866	.9854	.9863	.9861	.9860	.9859
1.2	.9853	9863	.9962	.9860	.9859	.9856	.9855	.9853	.9852	.9850	.9848	.9847	.9846
1.3	.9852	.9852	.9851	.9849	.9847	.9844	.9843	.9841	.9840	.9838	.9836	.9835	.9833
1.4	.9841	.9840	.9840	.9837	.9835	.9832	.9831	.9829	.9827	.9825	.9823	.9822	.9821
1.5	.9829	.9829	.9828	.9826	.9823	.9820	.9819	.9817	.9815	.9813	.9810	.9809	.9808
1.6	.9818	.9818	.9817	.9814	.9811	.9808	.9806	.9805	.9803	.9800	.9798	.9796	.9795
1.7	.9806	.9806	.9805	.9802	.9800	.9796	.9794	.9792	.9790	.9788	.9785	.9784	.9782
1.8	.9795	.9795	.9794	.9791	.9788	.9784	.9782	.9780	.9778	.9775	.9772	.9771	.9769
1.9	.9784	.9783	.9782	.9779	.9776	.9772	.9770	.9768	.9766	.9763	.9760	.9758	.9756
2.0	.9772	.9772	.9771	.9767	.9764	.9760	.9758	.9756	.9753	.9750	.9747	.9745	.9744
2.1	.9761	.9761	.9759	.9756	.9753	.9748	.9746	.9744	.9741	.9738	.9734	.9733	.9731
2.2	.9750	.9749	.9748	.9744	.9741	.9736	.9734	.9731	.9729	.9725	.9722	.9720	.9718
2.3	.9738	.9738	.9736	.9732	.9729	.9724	.9722	.9719	.9716	.9713	.9709	.9707	.9705
2.4	.9727	.9726	.9725	.9721	.9717	.9712	.9710	.9707	.9704	.9700	.9697	.9694	.9692
2.5	.9715	.9715	.9713	.9709	.9705	.9700	.9698	.9695	.9692	.9688	.9684	.9582	.9680
2.6	.9704	.9704	.9702	.9698	.9594	.9588	.9686	.9683	.9679	.9675	.9671	.9669	.9667
2.7	.9593	.9592	.9691	.9686	.9582	.9576	.9573	.9670	.9667	.9663	.9659	.9656	.9654
2.8	.9681	.9581	.9679	.9674	.9670	.9564	.9661	.9658	.9654	.9650	.9646	.9644	.9641
2.9	.9670	.9669	.9568	.9663	.9658	.9652	.9649	.9546	.9642	.9638	.9633	.9631	.9628
3.0	.9658	.9558	.9656	.9651	.9647	.9340	.9637	.9534	.9630	.9626	.9621	.9618	.9615
3.1	.9647	.9647	.9645	.9639	.9635	.9529	.9625	.9622	.9617	.9613	.9608	.9605	.9603
3.2	.9636	.9635	.9633	.9628	.9623	.9616	.9613	.9609	.9605	.9601	.9595	.9593	.9590
3.3	.9524	.9624	.9622	.9616	.9611	.9504	.9501	.9597	.9593	.9588	.9583	.9580	.9577
3.4	.9613	.9612	.9610	.9604	.9539	.9592	.9589	.9585	.9580	.9576	.9570	.9567	.9564
3.5	.9502	.9601	.9599	.9593	.9588	.9580	.9577	.9573	.9568	.9563	.9558	.9554	.9551
3.6	.9590	.9590	.9587	.9581	.9576	.9558	.9565	.9550	.9556	.9551	.9545	.9542	.9538
3.7	.9579	.9578	.9576	.9570	.9564	.9555	.9553	.9548	.9543	.9538	.9532	.9529	.9526
3.8	.9557	.9567	.9564	.9558	.9552	.9544	.9540	.9536	.9531	.9526	.9520	.9516	.9513
3.9	.9556	.9555	.9553	.9546	.9540	.9532	.9528	.9524	.9519	.9513	.9507	.9504	.9500
4.0	.9545	.9544	.9542	.9535	.9529	.9520	.9516	.9512	.9506	.9501	.9494	.9491	.9487

EXPANSION FACTORS — FLANGE TAPS — Y_1

Static Pressure Taken from Upstream Taps

$\frac{h_w}{P_{f1}}$ Ratio	$\beta = \frac{d}{D}$ Ratio												
	.63	.64	.65	.66	.67	.68	.69	.70	.71	.72	.73	.74	.75
0.0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	.10000
0.1	.9987	.9987	.9987	.9987	.9987	.9987	.9986	.9986	.9986	.9986	.9986	.9986	.9986
0.2	.9974	.9974	.9974	.9974	.9973	.9973	.9973	.9973	.9972	.9972	.9972	.9971	.9971
0.3	.9961	.9961	.9961	.9960	.9960	.9960	.9959	.9959	.9958	.9958	.9958	.9957	.9957
0.4	.9948	.9948	.9948	.9947	.9947	.9946	.9946	.9945	.9945	.9944	.9943	.9943	.9942
0.5	.9935	.9935	.9934	.9934	.9933	.9933	.9932	.9931	.9931	.9930	.9929	.9929	.9928
0.6	.9923	.9922	.9921	.9921	.9920	.9919	.9918	.9918	.9917	.9916	.9915	.9914	.9913
0.7	.9910	.9909	.9908	.9907	.9907	.9906	.9905	.9904	.9903	.9902	.9901	.9900	.9899
0.8	.9897	.9896	.9895	.9894	.9893	.9892	.9891	.9890	.9889	.9888	.9887	.9886	.9884
0.9	.9884	.9883	.9882	.9881	.9880	.9879	.9878	.9877	.9875	.9874	.9873	.9871	.9870
1.0	.9871	.9870	.9869	.9868	.9867	.9865	.9864	.9863	.9861	.9860	.9859	.9857	.9855
1.1	.9858	.9857	.9856	.9854	.9853	.9852	.9851	.9849	.9848	.9846	.9844	.9843	.9841
1.2	.9845	.9844	.9843	.9841	.9840	.9838	.9837	.9835	.9834	.9832	.9830	.9828	.9826
1.3	.9832	.9831	.9829	.9828	.9827	.9825	.9823	.9822	.9820	.9818	.9816	.9814	.9812
1.4	.9819	.9818	.9816	.9815	.9813	.9812	.9810	.9808	.9806	.9804	.9802	.9800	.9798
1.5	.9806	.9805	.9803	.9802	.9800	.9798	.9795	.9794	.9792	.9790	.9788	.9786	.9783
1.6	.9793	.9792	.9790	.9788	.9787	.9785	.9783	.9781	.9778	.9776	.9774	.9771	.9769
1.7	.9780	.9779	.9777	.9775	.9773	.9771	.9769	.9767	.9764	.9762	.9760	.9757	.9754
1.8	.9768	.9766	.9764	.9762	.9760	.9758	.9755	.9753	.9751	.9748	.9745	.9743	.9740
1.9	.9755	.9753	.9751	.9749	.9747	.9744	.9742	.9739	.9737	.9734	.9731	.9728	.9725
2.0	.9742	.9740	.9738	.9735	.9733	.9731	.9728	.9726	.9723	.9720	.9717	.9714	.9711
2.1	.9729	.9727	.9725	.9722	.9720	.9717	.9715	.9712	.9709	.9706	.9703	.9700	.9696
2.2	.9716	.9714	.9711	.9709	.9706	.9704	.9701	.9698	.9695	.9692	.9689	.9685	.9682
2.3	.9703	.9701	.9698	.9696	.9693	.9690	.9688	.9685	.9681	.9678	.9675	.9671	.9667
2.4	.9690	.9688	.9685	.9683	.9680	.9677	.9674	9671	.9668	.9664	.9661	.9657	.9653
2.5	.9677	.9675	.9672	.9669	.9666	.9663	.9660	.9657	.9654	.9650	.9646	.9643	.9639
2.6	.9664	.9662	.9659	.9656	.9653	.9650	.9647	.9643	.9640	.9636	.9632	.9628	.9624
2.7	.9651	.9649	.9646	.9643	.9640	.9637	.9633	.9630	.9626	.9622	.9618	.9614	.9610
2.8	.9638	.9635	.9633	.9630	.9626	.9623	.9620	.9616	.9612	.9608	.9604	.9600	.9595
2.9	.9625	.9623	.9620	.9616	.9613	.9610	.9606	.9602	.9598	.9594	.9590	.9585	.9581
3.0	.9613	.9610	.9606	.9603	.9600	.9596	.9592	.9588	.9584	.9580	.9576	.9571	.9566
3.1	.9600	.9597	.9593	.9590	.9586	.9583	.9579	.9575	.9571	.9566	.9562	.9557	.9552
3.2	.9587	.9584	.9580	.9577	.9573	.9569	.9565	.9561	.9557	.9552	.9547	.9542	.9537
3.3	.9574	.9571	.9567	.9564	.9560	.9555	.9552	.9547	.9543	.9538	.9533	.9528	.9523
3.4	.9561	.9558	.9554	.9550	.9546	.9542	.9538	.9534	.9529	.9524	.9519	.9514	.9508
3.5	.9548	.9545	.9541	.9537	.9533	.9529	.9524	.9520	.9515	.9510	.9505	.9500	.9494
3.6	.9535	.9532	.9528	.9524	.9520	.9515	.9511	.9506	.9501	.9496	.9491	.9485	.9480
3.7	.9522	.9518	.9515	.9511	.9506	.9502	.9497	.9492	.9487	.9482	.9477	.9471	.9465
3.8	.9509	.9505	.9502	.9497	.9493	.9488	.9484	.9479	.9474	.9468	.9463	.9457	.9451
3.9	.9496	.9492	.9488	.9484	.9480	.9475	.9470	.9465	.9460	.9454	.9448	.9442	.9436
4.0	.9483	.9479	.9475	.9471	.9466	.9462	.9457	.9451	.9446	.9440	.9434	.9428	.9422

EXPANSION FACTORS—FLANGE TAPS—Y_2

Static Pressure, Taken from Downstream Taps

$\frac{h_w}{p/z}$ Ratio	$\beta = \frac{d}{D}$ Ratio												
	.1	.2	.3	.4	.45	.50	.52	.54	.56	.58	.60	.61	.62
0.0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
0.1	1.0007	1.0007	1.0006	1.0006	1.0006	1.0006	1.0006	1.0006	1.0006	1.0006	1.0006	1.0005	1.0006
0.2	1.0013	1.0013	1.0013	1.0013	1.0012	1.0012	1.0012	1.0012	1.0011	1.0011	1.0011	1.0011	1.0010
0.3	1.0020	1.0020	1.0020	1.0019	1.0019	1.0018	1.0018	1.0018	1.0017	1.0017	1.0016	1.0016	1.0016
0.4	1.0027	1.0027	1.0026	1.0026	1.0025	1.0024	1.0024	1.0023	1.0023	1.0022	1.0022	1.0021	1.0021
0.5	1.0033	1.0033	1.0033	1.0032	1.0031	1.0030	1.0030	1.0029	1.0029	1.0028	1.0027	1.0027	1.0026
0.6	1.0040	1.0040	1.0040	1.0039	1.0038	1.0036	1.0036	1.0035	1.0034	1.0034	1.0033	1.0032	1.0032
0.7	1.0047	1.0047	1.0046	1.0045	1.0044	1.0043	1.0042	1.0041	1.0040	1.0039	1.0038	1.0038	1.0037
0.8	1.0054	1.0053	1.0053	1.0052	1.0050	1.0049	1.0048	1.0047	1.0046	1.0045	1.0044	1.0043	1.0042
0.9	1.0060	1.0060	1.0060	1.0058	1.0057	1.0055	1.0054	1.0053	1.0052	1.0050	1.0049	1.0048	1.0048
1.0	1.0067	1.0067	1.0066	1.0065	1.0063	1.0061	1.0060	1.0059	1.0058	1.0056	1.0055	1.0054	1.0053
1.1	1.0074	1.0074	1.0073	1.0071	1.0069	1.0067	1.0066	1.0065	1.0063	1.0062	1.0060	1.0059	1.0058
1.2	1.0080	1.0080	1.0080	1.0078	1.0076	1.0073	1.0072	1.0071	1.0069	1.0068	1.0066	1.0065	1.0064
1.3	1.0087	1.0087	1.0086	1.0084	1.0082	1.0080	1.0078	1.0077	1.0075	1.0073	1.0071	1.0070	1.0069
1.4	1.0094	1.0094	1.0093	1.0091	1.0089	1.0086	1.0084	1.0083	1.0081	1.0079	1.0077	1.0076	1.0074
1.5	1.0101	1.0101	1.0100	1.0097	1.0095	1.0092	1.0090	1.0089	1.0087	1.0085	1.0082	1.0081	1.0080
1.6	1.0108	1.0107	1.0106	1.0104	1.0101	1.0098	1.0096	1.0095	1.0093	1.0090	1.0088	1.0087	1.0085
1.7	1.0114	1.0114	1.0113	1.0110	1.0108	1.0104	1.0103	1.0101	1.0099	1.0096	1.0094	1.0092	1.0091
1.8	1.0121	1.0121	1.0120	1.0117	1.0114	1.0111	1.0109	1.0107	1.0104	1.0102	1.0099	1.0098	1.0096
1.9	1.0128	1.0128	1.0126	1.0123	1.0121	1.0117	1.0115	1.0113	1.0110	1.0108	1.0105	1.0103	1.0102
2.0	1.0135	1.0134	1.0133	1.0130	1.0127	1.0123	1.0121	1.0119	1.0116	1.0114	1.0110	1.0109	1.0107
2.1	1.0142	1.0141	1.0140	1.0136	1.0134	1.0129	1.0127	1.0125	1.0122	1.0119	1.0116	1.0114	1.0112
2.2	1.0148	1.0148	1.0147	1.0143	1.0140	1.0136	1.0133	1.0131	1.0128	1.0125	1.0122	1.0120	1.0118
2.3	1.0155	1.0155	1.0154	1.0150	1.0146	1.0142	1.0140	1.0137	1.0134	1.0131	1.0127	1.0126	1.0124
2.4	1.0162	1.0162	1.0160	1.0156	1.0153	1.0148	1.0146	1.0143	1.0140	1.0137	1.0133	1.0131	1.0129
2.5	1.0169	1.0168	1.0167	1.0163	1.0159	1.0154	1.0152	1.0149	1.0146	1.0142	1.0139	1.0137	1.0134
2.6	1.0176	1.0175	1.0174	1.0170	1.0166	1.0161	1.0158	1.0155	1.0152	1.0148	1.0144	1.0142	1.0140
2.7	1.0182	1.0182	1.0180	1.0176	1.0172	1.0167	1.0164	1.0161	1.0158	1.0154	1.0150	1.0148	1.0146
2.8	1.0189	1.0189	1.0187	1.0183	1.0179	1.0173	1.0170	1.0167	1.0164	1.0160	1.0156	1.0154	1.0151
2.9	1.0196	1.0196	1.0194	1.0189	1.0185	1.0180	1.0177	1.0173	1.0170	1.0166	1.0162	1.0159	1.0157
3.0	1.0203	1.0203	1.0201	1.0196	1.0192	1.0186	1.0183	1.0180	1.0176	1.0172	1.0167	1.0165	1.0162
3.1	1.0210	1.0210	1.0208	1.0203	1.0198	1.0192	1.0189	1.0186	1.0182	1.0178	1.0173	1.0170	1.0168
3.2	1.0217	1.0216	1.0214	1.0209	1.0205	1.0198	1.0195	1.0192	1.0188	1.0184	1.0179	1.0176	1.0173
3.3	1.0224	1.0223	1.0221	1.0216	1.0211	1.0205	1.0202	1.0198	1.0194	1.0189	1.0184	1.0182	1.0179
3.4	1.0230	1.0230	1.0228	1.0223	1.0218	1.0211	1.0208	1.0204	1.0200	1.0195	1.0190	1.0187	1.0184
3.5	1.0237	1.0237	1.0235	1.0229	1.0224	1.0217	1.0214	1.0210	1.0206	1.0201	1.0196	1.0193	1.0190
3.6	1.0244	1.0244	1.0242	1.0236	1.0231	1.0224	1.0220	1.0216	1.0212	1.0207	1.0202	1.0199	1.0196
3.7	1.0251	1.0251	1.0248	1.0243	1.0237	1.0230	1.0226	1.0222	1.0218	1.0213	1.0207	1.0204	1.0201
3.8	1.0258	1.0258	1.0255	1.0249	1.0244	1.0236	1.0233	1.0229	1.0224	1.0219	1.0213	1.0210	1.0207
3.9	1.0265	1.0264	1.0262	1.0256	1.0250	1.0243	1.0239	1.0235	1.0230	1.0225	1.0219	1.0216	1.0213
4.0	1.0272	1.0271	1.0269	1.0263	1.0257	1.0249	1.0246	1.0241	1.0236	1.0231	1.0225	1.0222	1.0218

EXPANSION FACTORS—FLANGE TAPS—Y_2

Static Pressure, Taken from Downstream Taps

$\frac{h_w}{p/z}$ Ratio	$\beta = \frac{d}{D}$ Ratio												
	.63	.64	.65	.66	.67	.68	.69	.70	.71	.72	.73	.74	.75
0.0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
0.1	1.0005	1.0005	1.0005	1.0005	1.0005	1.0004	1.0004	1.0004	1.0004	1.0004	1.0004	1.0004	1.0004
0.2	1.0010	1.0010	1.0010	1.0010	1.0009	1.0009	1.0009	1.0009	1.0008	1.0008	1.0008	1.0008	1.0007
0.3	1.0015	1.0015	1.0015	1.0014	1.0014	1.0014	1.0013	1.0013	1.0013	1.0012	1.0012	1.0011	1.0011
0.4	1.0021	1.0020	1.0020	1.0019	1.0019	1.0018	1.0018	1.0017	1.0017	1.0016	1.0016	1.0015	1.0014
0.5	1.0026	1.0025	1.0025	1.0024	1.0024	1.0023	1.0022	1.0022	1.0021	1.0020	1.0020	1.0019	1.0018
0.6	1.0031	1.0030	1.0030	1.0029	1.0028	1.0028	1.0027	1.0026	1.0025	1.0025	1.0024	1.0023	1.0022
0.7	1.0036	1.0036	1.0035	1.0034	1.0033	1.0032	1.0032	1.0031	1.0030	1.0029	1.0028	1.0027	1.0026
0.8	1.0042	1.0041	1.0040	1.0039	1.0038	1.0037	1.0036	1.0035	1.0034	1.0033	1.0032	1.0030	1.0029
0.9	1.0047	1.0046	1.0045	1.0044	1.0043	1.0042	1.0041	1.0040	1.0038	1.0037	1.0036	1.0034	1.0033
1.0	1.0052	1.0051	1.0050	1.0049	1.0048	1.0047	1.0045	1.0044	1.0043	1.0041	1.0040	1.0038	1.0037
1.1	1.0057	1.0056	1.0055	1.0054	1.0053	1.0051	1.0050	1.0049	1.0047	1.0046	1.0044	1.0042	1.0041
1.2	1.0062	1.0061	1.0060	1.0059	1.0058	1.0056	1.0055	1.0053	1.0052	1.0050	1.0048	1.0046	1.0044
1.3	1.0068	1.0066	1.0065	1.0064	1.0062	1.0061	1.0059	1.0058	1.0056	1.0054	1.0052	1.0050	1.0048
1.4	1.0073	1.0072	1.0070	1.0069	1.0067	1.0066	1.0064	1.0062	1.0060	1.0058	1.0056	1.0054	1.0052
1.5	1.0078	1.0077	1.0076	1.0074	1.0072	1.0070	1.0069	1.0067	1.0065	1.0063	1.0060	1.0058	1.0056
1.6	1.0084	1.0082	1.0081	1.0079	1.0077	1.0075	1.0073	1.0071	1.0069	1.0067	1.0065	1.0062	1.0060
1.7	1.0089	1.0088	1.0086	1.0084	1.0082	1.0080	1.0078	1.0076	1.0074	1.0071	1.0069	1.0066	1.0064
1.8	1.0094	1.0093	1.0091	1.0089	1.0087	1.0085	1.0083	1.0080	1.0078	1.0076	1.0073	1.0070	1.0068
1.9	1.0100	1.0098	1.0096	1.0094	1.0092	1.0090	1.0088	1.0085	1.0083	1.0080	1.0077	1.0074	1.0071
2.0	1.0105	1.0103	1.0101	1.0099	1.0097	1.0095	1.0092	1.0090	1.0087	1.0084	1.0081	1.0078	1.0075
2.1	1.0111	1.0109	1.0106	1.0104	1.0102	1.0100	1.0097	1.0094	1.0092	1.0089	1.0086	1.0083	1.0079
2.2	1.0116	1.0114	1.0112	1.0109	1.0107	1.0104	1.0102	1.0099	1.0096	1.0093	1.0090	1.0087	1.0083
2.3	1.0121	1.0119	1.0117	1.0114	1.0112	1.0109	1.0106	1.0104	1.0101	1.0098	1.0094	1.0091	1.0087
2.4	1.0127	1.0124	1.0122	1.0120	1.0117	1.0114	1.0111	1.0108	1.0105	1.0102	1.0098	1.0095	1.0091
2.5	1.0132	1.0130	1.0127	1.0125	1.0122	1.0119	1.0116	1.0113	1.0110	1.0106	1.0103	1.0099	1.0095
2.6	1.0138	1.0135	1.0133	1.0130	1.0127	1.0124	1.0121	1.0118	1.0114	1.0111	1.0107	1.0103	1.0099
2.7	1.0143	1.0140	1.0138	1.0135	1.0132	1.0129	1.0126	1.0122	1.0119	1.0115	1.0111	1.0107	1.0103
2.8	1.0148	1.0146	1.0143	1.0140	1.0137	1.0134	1.0131	1.0127	1.0124	1.0120	1.0116	1.0112	1.0107
2.9	1.0154	1.0151	1.0148	1.0145	1.0142	1.0139	1.0136	1.0132	1.0128	1.0124	1.0120	1.0116	1.0111
3.0	1.0160	1.0157	1.0154	1.0150	1.0147	1.0144	1.0140	1.0137	1.0133	1.0129	1.0124	1.0120	1.0116
3.1	1.0165	1.0162	1.0159	1.0156	1.0152	1.0149	1.0145	1.0141	1.0137	1.0133	1.0129	1.0124	1.0120
3.2	1.0170	1.0167	1.0164	1.0161	1.0158	1.0154	1.0150	1.0146	1.0142	1.0138	1.0133	1.0128	1.0124
3.3	1.0176	1.0173	1.0170	1.0166	1.0163	1.0159	1.0156	1.0151	1.0147	1.0142	1.0138	1.0133	1.0128
3.4	1.0181	1.0178	1.0175	1.0171	1.0168	1.0164	1.0160	1.0156	1.0151	1.0147	1.0142	1.0137	1.0132
3.5	1.0187	1.0184	1.0180	1.0177	1.0173	1.0169	1.0165	1.0160	1.0156	1.0151	1.0146	1.0141	1.0136
3.6	1.0192	1.0189	1.0186	1.0182	1.0178	1.0174	1.0170	1.0165	1.0161	1.0156	1.0151	1.0146	1.0140
3.7	1.0198	1.0195	1.0191	1.0187	1.0183	1.0179	1.0175	1.0170	1.0165	1.0160	1.0156	1.0150	1.0144
3.8	1.0204	1.0200	1.0196	1.0192	1.0188	1.0184	1.0180	1.0175	1.0170	1.0165	1.0160	1.0154	1.0148
3.9	1.0209	1.0206	1.0202	1.0198	1.0194	1.0189	1.0185	1.0180	1.0175	1.0170	1.0164	1.0159	1.0153
4.0	1.0215	1.0211	1.0207	1.0203	1.0199	1.0194	1.0190	1.0185	1.0180	1.0174	1.0169	1.0163	1.0157

OILS, SPECIFIC GRAVITY—TEMPERATURE FACTORS—F_{gt}
BELLOWS TYPE METERS

$$F_{gt} = \frac{1.0057}{\sqrt{G_1}} \sqrt{\frac{G_f}{G_1}}$$

Degrees A.P.I. at 60°F	Liquid specific gravity at 60°F	Temperature of flowing oil, °F.										
		0	20	40	60	80	100	125	150	200	250	300
151.5	.50	1.4885	1.4669	1.4449	1.4223	1.3991	1.3753	1.3446	1.3129	1.2459	1.1735	1.0946
145.9	.51	1.4707	1.4504	1.4296	1.4083	1.3864	1.3640	1.3352	1.3054	1.2426	1.1750	1.1017
140.6	.52	1.4535	1.4344	1.4147	1.3947	1.3741	1.3530	1.3259	1.2980	1.2391	1.1760	1.1078
135.5	.53	1.4369	1.4189	1.4004	1.3814	1.3621	1.3423	1.3168	1.2906	1.2355	1.1765	1.1131
130.5	.54	1.4209	1.4038	1.3864	1.3686	1.3504	1.3317	1.3078	1.2832	1.2316	1.1766	1.1175
125.8	.55	1.4053	1.3892	1.3728	1.3561	1.3390	1.3215	1.2990	1.2759	1.2276	1.1762	1.1213
121.2	.56	1.3902	1.3751	1.3597	1.3439	1.3278	1.3114	1.2903	1.2687	1.2235	1.1755	1.1244
116.7	.57	1.3756	1.3614	1.3469	1.3321	1.3170	1.3016	1.2818	1.2615	1.2193	1.1745	1.1270
112.5	.58	1.3616	1.3482	1.3345	1.3205	1.3063	1.2918	1.2732	1.2542	1.2145	1.1726	1.1282
108.3	.59	1.3481	1.3354	1.3225	1.3093	1.2959	1.2822	1.2647	1.2468	1.2095	1.1702	1.1286
104.3	.60	1.3350	1.3230	1.3108	1.2984	1.2857	1.2727	1.2562	1.2393	1.2042	1.1672	1.1282
100.5	.61	1.3227	1.3112	1.2995	1.2877	1.2756	1.2632	1.2475	1.2314	1.1981	1.1630	1.1261
96.7	.62	1.3107	1.2997	1.2886	1.2772	1.2657	1.2540	1.2390	1.2237	1.1920	1.1587	1.1238
93.1	.63	1.2991	1.2886	1.2779	1.2671	1.2560	1.2448	1.2305	1.2159	1.1857	1.1540	1.1208
89.8	.64	1.2878	1.2777	1.2675	1.2571	1.2466	1.2358	1.2222	1.2082	1.1794	1.1493	1.1178
86.2	.65	1.2769	1.2672	1.2574	1.2474	1.2373	1.2270	1.2138	1.2005	1.1728	1.1440	1.1140
82.9	.66	1.2663	1.2570	1.2475	1.2379	1.2282	1.2182	1.2057	1.1928	1.1664	1.1388	1.1101
79.7	.67	1.2560	1.2470	1.2379	1.2287	1.2193	1.2097	1.1976	1.1853	1.1599	1.1335	1.1060
76.6	.68	1.2460	1.2373	1.2285	1.2196	1.2105	1.2014	1.1897	1.1778	1.1535	1.1281	1.1018
73.6	.69	1.2362	1.2278	1.2193	1.2107	1.2020	1.1932	1.1819	1.1705	1.1471	1.1228	1.0976
70.6	.70	1.2267	1.2186	1.2104	1.2020	1.1936	1.1851	1.1743	1.1633	1.1408	1.1174	1.0933
67.8	.71	1.2173	1.2095	1.2016	1.1935	1.1854	1.1772	1.1668	1.1562	1.1346	1.1122	1.0891
65.0	.72	1.2082	1.2006	1.1930	1.1852	1.1774	1.1696	1.1595	1.1493	1.1285	1.1071	1.0850
62.3	.73	1.1992	1.1919	1.1845	1.1771	1.1696	1.1619	1.1523	1.1425	1.1225	1.1020	1.0808
59.7	.74	1.1905	1.1834	1.1763	1.1691	1.1618	1.1545	1.1452	1.1358	1.1166	1.0969	1.0766
57.2	.75	1.1819	1.1751	1.1682	1.1613	1.1543	1.1472	1.1383	1.1292	1.1108	1.0919	1.0725
54.7	.76	1.1734	1.1669	1.1603	1.1536	1.1469	1.1401	1.1316	1.1229	1.1053	1.0873	1.0689
52.3	.77	1.1650	1.1588	1.1525	1.1461	1.1397	1.1332	1.1250	1.1168	1.1000	1.0829	1.0653
49.9	.78	1.1570	1.1509	1.1449	1.1387	1.1326	1.1263	1.1185	1.1106	1.0945	1.0782	1.0614
47.6	.79	1.1490	1.1432	1.1374	1.1315	1.1256	1.1196	1.1121	1.1045	1.0891	1.0735	1.0575
45.4	.80	1.1413	1.1357	1.1301	1.1244	1.1187	1.1130	1.1057	1.0984	1.0837	1.0687	1.0535
43.2	.81	1.1338	1.1284	1.1229	1.1174	1.1119	1.1064	1.0994	1.0924	1.0783	1.0639	1.0493
41.1	.82	1.1264	1.1212	1.1159	1.1106	1.1053	1.0999	1.0932	1.0864	1.0728	1.0590	1.0450
39.0	.83	1.1193	1.1142	1.1090	1.1039	1.0987	1.0936	1.0871	1.0805	1.0674	1.0541	1.0406
36.9	.84	1.1122	1.1073	1.1023	1.0973	1.0923	1.0873	1.0810	1.0746	1.0619	1.0490	1.0361
35.0	.85	1.1054	1.1005	1.0957	1.0908	1.0860	1.0811	1.0750	1.0688	1.0565	1.0441	1.0316
33.0	.86	1.0987	1.0939	1.0892	1.0845	1.0797	1.0750	1.0690	1.0630	1.0511	1.0390	1.0270
31.1	.87	1.0921	1.0875	1.0828	1.0782	1.0736	1.0690	1.0632	1.0574	1.0457	1.0341	1.0224
29.3	.88	1.0856	1.0811	1.0766	1.0721	1.0676	1.0630	1.0574	1.0517	1.0404	1.0291	1.0178
27.5	.89	1.0793	1.0749	1.0705	1.0660	1.0616	1.0572	1.0517	1.0461	1.0351	1.0241	1.0131
25.7	.90	1.0731	1.0688	1.0644	1.0601	1.0558	1.0514	1.0460	1.0407	1.0299	1.0192	1.0085
24.0	.91	1.0671	1.0628	1.0585	1.0543	1.0500	1.0458	1.0405	1.0352	1.0247	1.0142	1.0038
22.3	.92	1.0611	1.0569	1.0527	1.0485	1.0443	1.0402	1.0350	1.0298	1.0195	1.0093	0.9992
20.7	.93	1.0553	1.0511	1.0470	1.0429	1.0388	1.0347	1.0296	1.0245	1.0144	1.0044	0.9946
19.0	.94	1.0495	1.0454	1.0414	1.0373	1.0333	1.0292	1.0242	1.0193	1.0094	0.9997	0.9900
17.5	.95	1.0438	1.0398	1.0358	1.0318	1.0279	1.0239	1.0190	1.0141	1.0045	0.9949	0.9855
15.9	.96	1.0383	1.0343	1.0304	1.0264	1.0225	1.0187	1.0138	1.0091	0.9996	0.9903	0.9811
14.4	.97	1.0328	1.0289	1.0250	1.0211	1.0173	1.0135	1.0088	1.0041	0.9948	0.9857	0.9768
12.9	.98	1.0274	1.0235	1.0197	1.0159	1.0121	1.0084	1.0037	0.9991	0.9901	0.9812	0.9725
11.4	.99	1.0221	1.0183	1.0145	1.0108	1.0071	1.0034	0.9988	0.9943	0.9854	0.9767	0.9682
10.0	1.00	1.0168	1.0131	1.0094	1.0057	1.0020	0.9984	0.9939	0.9895	0.9808	0.9723	0.9640
8.6	1.01	1.0116	1.0079	1.0043	1.0007	0.9971	0.9936	0.9891	0.9847	0.9761	0.9676	0.9592
7.2	1.02	1.0064	1.0029	0.9993	0.9958	0.9923	0.9888	0.9844	0.9801	0.9714	0.9629	0.9545
5.8	1.03	1.0013	0.9979	0.9944	0.9909	0.9875	0.9840	0.9797	0.9754	0.9669	0.9583	0.9498
4.6	1.04	0.9963	0.9929	0.9896	0.9862	0.9828	0.9794	0.9751	0.9709	0.9623	0.9538	0.9452

FLOWING TEMPERATURE FACTORS—F_{tf}

$$F_{tf} = \sqrt{\frac{520}{460 + \text{actual flowing temperature}}}$$

°F.	Factor	°F.	Factor	°F.	Factor	°F.	Factor	°F.	Factor	°F.	Factor
1	1.0621	21	1.0398	41	1.0188	61	0.9990	81	0.9804	110	0.9551
2	1.0609	22	1.0387	42	1.0178	62	0.9981	82	0.9795	120	0.9469
3	1.0598	23	1.0376	43	1.0168	63	0.9971	83	0.9786	130	0.9388
4	1.0586	24	1.0365	44	1.0157	64	0.9962	84	0.9777	140	0.9309
5	1.0575	25	1.0355	45	1.0147	65	0.9952	85	0.9768	150	0.9233
6	1.0564	26	1.0344	46	1.0137	66	0.9943	86	0.9759	160	0.9158
7	1.0552	27	1.0333	47	1.0127	67	0.9933	87	0.9750	170	0.9085
8	1.0541	28	1.0323	48	1.0117	68	0.9924	88	0.9741	180	0.9014
9	1.0530	29	1.0312	49	1.0107	69	0.9915	89	0.9732	190	0.8944
10	1.0518	30	1.0302	50	1.0098	70	0.9905	90	0.9723	200	0.8876
11	1.0507	31	1.0291	51	1.0088	71	0.9896	91	0.9715	210	0.8810
12	1.0496	32	1.0281	52	1.0078	72	0.9887	92	0.9706	220	0.8745
13	1.0485	33	1.0270	53	1.0068	73	0.9877	93	0.9697	230	0.8681
14	1.0474	34	1.0260	54	1.0058	74	0.9868	94	0.9688	240	0.8619
15	1.0463	35	1.0249	55	1.0048	75	0.9859	95	0.9680	250	0.8558
16	1.0452	36	1.0239	56	1.0039	76	0.9850	96	0.9671	260	0.8498
17	1.0441	37	1.0229	57	1.0029	77	0.9840	97	0.9662	270	0.8440
18	1.0430	38	1.0219	58	1.0019	78	0.9831	98	0.9653	280	0.8383
19	1.0419	39	1.0208	59	1.0010	79	0.9822	99	0.9645	290	0.8327
20	1.0408	40	1.0198	60	1.0000	80	0.9813	100	0.9636	300	0.8272

PRESSURE BASE FACTORS—F_{pb}

F_{pb} = 14.73 + base pressure, psia

Pressure base, psia	Factor F_{pb}	Pressure base, psia	Factor F_{pb}
14.4	1.0229	15.025 (10 oz. above 14.4)	0.9804
14.65 (4 oz. above 14.4)	1.0055	15.2 (8 oz. above 14.7)	0.9691
14.696	1.0023	15.325 (10 oz. above 14.7)	0.9612
14.73	1.0000	15.4 (1 psi above 14.4)	0.9565
14.735	0.9997	15.7 (1 psi above 14.7)	0.9382
14.9 (8 oz. above 14.4)	0.9886	16.4 (2 psi above 14.4)	0.8982
14.95 (4 oz. above 14.7)	0.9853	16.7 (2 psi above 14.7)	0.8820

TEMPERATURE BASE FACTORS—F_{tb}

$$F_{tb} = \frac{460 + \text{Temperature base °F}}{520}$$

Temperature base °F	Factor F_{tb}	Temperature base °F	Factor F_{tb}	Temperature base °F	Factor F_{tb}
45	0.9712	65	1.0096	85	1.0481
50	0.9808	70	1.0192	90	1.0577
55	0.9904	75	1.0288	95	1.0673
60	1.0000	80	1.0385	100	1.0769

$F_s = \dfrac{1.0618}{\sqrt{p_f{}^s}}$, For *Bellows-Type* Meters

Pressure, psig	Total Steam Temperature, °F										
	300	350	400	450	500	550	600	700	800	1000	1200
25	.0505	.0488	.0472	.0459	.0446	.0435	.0424	.0405	.0388	.0360	.0338
50	.0510	.0491	.0475	.0460	.0447	.0436	.0425	.0405	.0389	.0361	.0338
75		.0494	.0477	.0462	.0449	.0437	.0426	.0406	.0389	.0361	.0338
100		.0498	.0479	.0464	.0450	.0438	.0426	.0407	.0389	.0361	.0338
125		.0501	.0482	.0466	.0451	.0439	.0427	.0407	.0390	.0361	.0338
150			.0485	.0468	.0453	.0440	.0428	.0408	.0390	.0361	.0338
175			.0488	.0470	.0454	.0441	.0429	.0408	.0391	.0362	.0339
200			.0491	.0472	.0456	.0442	.0430	.0409	.0391	.0362	.0339
250				.0476	.0459	.0444	.0432	.0410	.0392	.0362	.0339
300				.0481	.0462	.0447	.0434	.0411	.0393	.0363	.0339
350				.0486	.0466	.0449	.0436	.0412	.0393	.0363	.0339
400				.0492	.0470	.0452	.0438	.0414	.0394	.0364	.0340
450					.0474	.0455	.0440	.0415	.0395	.0364	.0340
500					.0478	.0458	.0442	.0416	.0396	.0364	.0340
600					.0488	.0465	.0446	.0419	.0398	.0365	.0341
700						.0472	.0451	.0422	.0400	.0366	.0341
800						.0480	.0457	.0425	.0401	.0367	.0341
900						.0489	.0463	.0428	.0403	.0368	.0342
1000						.0500	.0469	.0431	.0405	.0369	.0342
1100							.0477	.0435	.0407	.0370	.0343
1200							.0485	.0438	.0409	.0371	.0343
1300							.0494	.0442	.0412	.0372	.0344
1400							.0505	.0446	.0414	.0372	.0344
1500							.0518	.0450	.0416	.0373	.0345

$F_s = \dfrac{1.0618}{\sqrt{p_f{}^s}}$, For *Bellows-Type* Meters

Pressure, psig	Steam Quality, Per Cent										
	100	99	98	97	96	95	94	93	92	91	90
1	.0534	.0537	.0539	.0542	.0545	.0548	.0551	.0554	.0557	.0560	.0563
5	.0530	.0533	.0535	.0538	.0541	.0544	.0547	.0550	.0553	.0556	.0559
10	.0526	.0529	.0531	.0534	.0537	.0540	.0543	.0546	.0549	.0552	.0555
15	.0523	.0526	.0528	.0531	.0534	.0537	.0539	.0542	.0545	.0548	.0551
20	.0520	.0523	.0526	.0528	.0531	.0534	.0537	.0540	.0543	.0546	.0549
25	.0518	.0521	.0524	.0526	.0529	.0532	.0535	.0537	.0540	.0543	.0546
50	.0511	.0513	.0516	.0518	.0521	.0524	.0527	.0529	.0532	.0535	.0538
75	.0506	.0508	.0511	.0514	.0516	.0519	.0522	.0525	.0527	.0530	.0533
100	.0503	.0505	.0508	.0510	.0513	.0516	.0518	.0521	.0524	.0527	.0530
125	.0500	.0503	.0505	.0508	.0510	.0513	.0516	.0519	.0521	.0524	.0527
150	.0498	.0501	.0503	.0506	.0509	.0511	.0514	.0517	.0519	.0522	.0525
175	.0497	.0499	.0502	.0504	.0507	.0510	.0512	.0515	.0518	.0521	.0524
200	.0496	.0498	.0501	.0503	.0506	.0509	.0511	.0514	.0517	.0520	.0522
250	.0494	.0497	.0499	.0502	.0504	.0507	.0510	.0512	.0515	.0518	.0521
300	.0493	.0496	.0498	.0501	.0503	.0506	.0509	.0511	.0514	.0517	.0520
350	.0493	.0496	.0498	.0500	.0503	.0505	.0508	.0511	.0513	.0516	.0519
400	.0493	.0495	.0498	.0500	.0503	.0505	.0508	.0511	.0513	.0516	.0519
450	.0493	.0495	.0498	.0500	.0503	.0505	.0508	.0511	.0513	.0516	.0519
500	.0493	.0496	.0498	.0501	.0503	.0506	.0508	.0511	.0514	.0516	.0519
550	.0494	.0496	.0499	.0501	.0504	.0506	.0509	.0511	.0514	.0517	.0520
600	.0494	.0497	.0499	.0502	.0504	.0507	.0509	.0512	.0515	.0517	.0520

APPENDIX A

SAMPLING

Reliable sampling is essential to the proper specification and operation of production systems and plants. Although the discussion which follows is directed toward fluid sampling to obtain composition, remember that this is useful only when accompanied by accurate measurements of pressure, temperature and flow rate.

Samples are taken at different times and for different purposes.

1. In the exploratory stage before a production system is in place and regular production has been initiated.
2. After initiation of production but before basic development is complete.
3. The period of stable production (relatively speaking).
4. System and plant surveillance tests to merely monitor operations or obtain feed-back.
5. Acceptance tests.

Sampling is a simple process mechanically, and therein lies the danger. The average feeling is that "any fool can grab a sample." This is true, but what is needed is a representative sample. Consequently, sampling is fraught with pitfalls, the chief of which are:

1. Use of dirty cylinders that contain portions of past fluids analyzed.
2. Sampling at the wrong point to get a representative sample.
3. Obtaining sample when temperature and pressure conditions are not stable.
4. Taking the sample in such a manner that the pressure in the sampling cylinder is not the same as that in the separator.
5. Dilution of a gas sample with air.
6. Loss of sampling cylinder pressure between the time of sampling and analysis.
7. Failure to record all necessary data and properly mark the sample cylinders.

Any or all of these may introduce serious error. Many embarrassing and costly errors have resulted from improper sampling. Even under ideal sampling conditions it must be recognized that well-stream analyses are constantly varying over short periods of time and gradually vary, of course, during reservoir depletion.

SAMPLING TIME

Early Reservoir Fluid Samples

In frontier and offshore environments it is common to drill exploratory wells that will not be completed as producers. One purpose is to obtain fluid samples. Said samples will be the sole basis for the design of equipment for exploration of said reservoir (if deemed commercial) by virtue of the time-frame involved.

At best, the quality of these fluid samples is mediocre. Drill-stem test samples are basically worse than those taken using wire line sampling procedures. The basic problem in either case is the fact that an equilibrium sample is not obtained, particularly if both vapor and liquid are involved. But, this may be the only sample you have to design and specify the system. There are two basic choices: (1) "play like" the sample is absolutely correct and use it as such, or (2) accept the uncertainty and allow for it in the engineering exercise which follows. However imperfectly the latter is done, it always is the best choice.

At this stage some basic geological input is needed. How many likely producing zones are there? What is a reasonable guess about the number and location of wells? What reservoir water and solids problems may be encountered? At least 15-20 such pertinent questions are *critical* to design and specification. From a sampling viewpoint, a sample is needed from each producing horizon. Then a scenario must be developed about how these may be co-mingled during production.

Some reservoir engineering input is required because the present sample – no matter how good – only represents present behavior. Some extrapolation is required to predict future performance. What effect will this have on the change in magnitude and direction of current analyses?

Stable Reservoir Fluid Samples

Once the reservoir is produced at a constant rate for a reasonable period of time (usually a minimum of 72 hours) it is possible to obtain an equilibrium sample. This can be the first truly good sample available. The basic production system already may be in place but it is time to rethink earlier decisions and plan for the future changes that will occur inevitably.

Sampling at this stage can be very reliable if provisions were made in the design for P, T or fluid sampling points, and corresponding flow measurements. At this point in time the best reservoir fluid analyses will be obtained through use of a properly instrumented test separator.

Plant and System Fluid Samples

The data required to design equipment are the same data needed to monitor it. The mechanical designer must understand what is needed so the necessary sampling points are provided. If this is not done some very "silly situations" can arise. A common one is buying equipment based on a vendor's guarantee and then not providing the means to check said guarantee.

A critical aspect is preliminary planning of the test (and sampling). What is the purpose of the exercise? How many samples should be taken, where and at what frequency? What components need analyzing? What size samples are required? Sampling should not proceed until questions like these are answered.

Consider the time between sampling and analysis. A gas containing trace amounts of sulfur may test sweet if too long a time occurs because the minute (but important) quantity available has reacted with the alloy cylinder being used. Certain delayed tests on water are virtually meaningless.

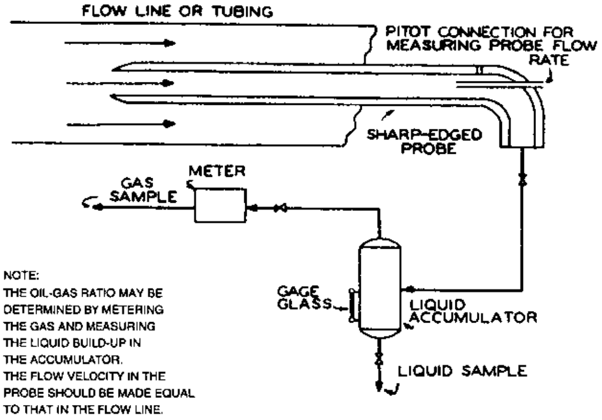
TWO-PHASE SAMPLING APPROACH

Most well streams, before separation, consist of a mixture of gas and liquid in varying proportions. This may vary from a dry-gas stream containing virtually no liquid to a crude-oil well with a very low gas-oil ratio. It is possible to obtain an analysis of these mixed streams at a point between the wellhead and the first separator.

Stinger

Figure A.1 shows a method for attempting to obtain a representative sample of a two-phase stream.

A knife-edged probe is inserted into the well stream, usually at the center line. It is sized and adjusted so that the velocity in the probe is equal to that in the pipe. Consequently, it is expected that over a reasonable period of time a representative sample will be obtained. When carefully done this method may be satisfactory but does have the following drawbacks:



NOTE:
THE OIL-GAS RATIO MAY BE DETERMINED BY METERING THE GAS AND MEASURING THE LIQUID BUILD-UP IN THE ACCUMULATOR.
THE FLOW VELOCITY IN THE PROBE SHOULD BE MADE EQUAL TO THAT IN THE FLOW LINE.

Figure A.1 Use of a "Stinger" for Sampling

1. When only small amounts of liquid are present or the velocity is low, or both, the liquid tends to flow along the walls.
2. Many wells, particularly gas-condensate, tend to produce liquid in surges so that a relatively long sampling period is needed. During such surges it is difficult to obtain representative samples.
3. Good equipment, and a trained operator, are needed for good results.

Separation Vessels

The most reliable method for sampling two-phase fluids is use of a test separator. As with a "stinger" the well must have been flowed at constant rate long enough so that conditions have stabilized. The time varies with the reservoir and flow rate but any time less than 72 hours is suspect.

A special test separator will have proper meters and sampling points built into the system. The gas and liquid flow rates, and analyses, will be re-combined to determine composite wellstream analysis as shown in Chapter 5.

A regular separator, without a liquid meter, may be used for this purpose by using level change in the gauge glass as a measure of liquid production. Based on vessel internal diameter, a given change in level represents a certain volume accumulation. Figure A.2 shows one possible hookup.

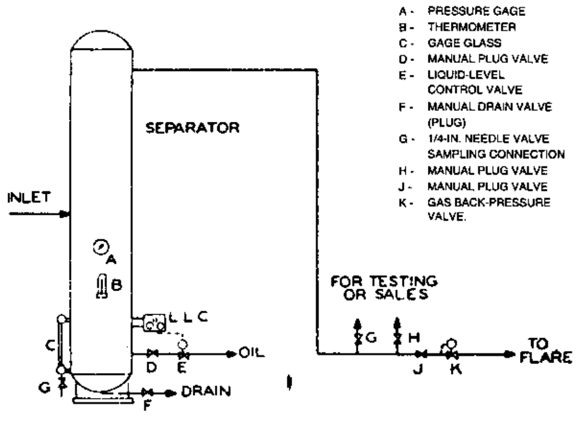


Figure A.2 Recommended Separator Hookup

Sampling locations are provided on the gauge and on the overhead gas line. The plug valve D is provided to have a positive shutoff ahead of the liquid level valve E, the drain also being likewise equipped (valve F). Except for valves D, F, and G, all accessories shown are standard with separator manufactures.

The method of handling the overhead gas is recommended whether the gas is being flared or sold. Valve J may be omitted, if desired, although it is handy in the event the backpressure valve fails to function.

If the gas is being flared, connection H may be used to test the gas volume by closing off valves J or K, or both, and opening H. An orifice well tester or critical flow prover may be simply connected to the outlet of valve H.

If the gas is being sold through connection H the backpressure valve will still keep the separator pressure from going too high by flaring any excess gas automatically. In such cases it is often wise to also have a back-pressure controller in the sales line to keep from drawing down the separator pressure. This is particularly true when a group of low-pressure separators are tied together on the suction side of a compressor station.

Sampling a stream at a separator is the recommended procedure, for it provides control over all the factors. When sampling in this manner it is necessary to determine the gas-oil ratio in the separator as well as obtaining the liquid and gas samples.

The gas sample should be taken close to the separator but after the gas has left the separator proper. It is inadvisable to draw it from the separator shell, for the walls of that vessel are covered with liquid which will contaminate the sample.

A very small amount of liquid will cause an appreciable error in a gas sample. For these reasons, the gas sample is normally obtained from a meter run (if present) or from connections G and H.

When no such connections are available it is sometimes absolutely necessary to obtain the gas sample from the pressure-gauge connection A in the shell of the separator. In this case it is necessary to improvise a method of scrubbing out any entrained liquid. On low-pressure separators an instrument gas drip pot may be used. A length of pipe filled with desiccant is also good. Some similar piece of equipment is a recommended part of the sampler's kit of tools.

The liquid samples may be normally withdrawn from the separator gauge glass. Installation of a 1/4 in. needle valve (G) is recommended for this purpose.

Any separation vessel possessing a liquid level control can at least be sampled for liquid. However, there may not be a liquid metering facility. One alternative approach is to close off the liquid dump and measure the time it takes the liquid level to rise between two marks on the gauge glass. Knowing the vessel size, one can estimate the liquid flow rate at that point in time.

Sometimes no gas meter is available for that specific vessel but one can calculate a reasonable quantity. It is a good idea though to provide at least an orifice flange or turbine meter connections even though they are in service only during testing. Sometimes a by-pass metering loop, ordinarily not in service, is a good idea.

METHOD OF SAMPLING

There are two basic methods of filling a sample bomb: (1) displacement of a liquid, or (2) use evacuated bombs. Both may be satisfactory if properly done.

Water is a convenient liquid because hydrocarbons are almost totally insoluble in it. If very water soluble materials like hydrogen sulfide and carbon dioxide are present water displacement is not suitable. For sampling cold liquids below the freezing point of water a 50-50 mixture with a glycol may be used.

Many prefer liquid displacement for it provides a positive means to exclude air from the sample. One furthermore has a positive indication of the rate of sampling. A sample can be taken at full system pressure.

The major problem with evacuated sample bombs is possible air leakage. Initially at least, the bomb will be below system pressure and some phase change is involved.

SAMPLE CONTAINER CONSTRUCTION

All sample containers used, regardless of the pressure rating, should be cylinders with valves on both ends. The ends should be closed either by (a) swages reducing to 1/4 in. diameter, threaded 1/4 in., or (b) by pipe caps which have been drilled in the center to accommodate 1/4 in. nipples, welded to the caps.

Service	Diameter (in. O.D.)	Pressure range (psig)
Gas	2. 3/8	Above 225
Gas	3. 1/2	90-225
Gas	12.	15-90
Gas	6.	30-400
Liquid	2. 3/8	(†)
Gas	(‡)	0-15

† 0 to working pressure of container
‡ 5 gallon water bottle

All fabricated containers should be made of extra-heavy seamless pipe, using 1/4 x 3 in. extra-heavy nipples and barstock needle valves. The working pressure should be stamped on the barrel of the container. In every case the container should be hydrostatically tested to 1 1/2 times the stamped working pressure.

Most metal sample containers are made of alloy steel, usually some form of stainless steel. Such alloys are used because many fluids contain at least trace quantities of sulfur compounds. Sulfur compounds do react with ordinary stainless steel to some degree. If only trace amounts are present all of it may react with the metal wall before analysis which leads to erroneous conclusions. New alloys are being used to minimize this.

It is advisable that a given sample container not be used for sampling both gas and liquid, to prevent undue contamination. To avoid mixup, gas sample containers should be painted aluminum between the seams, with the ends painted red. They should have the following stenciled in red on the shell:

Gas: Max. W. P. = XXXX bars [psig]

Liquids containers should be painted red only, with the following stenciled in aluminum on the shell:

Liquid: Max. W. P. = XXXX bars [psig]

It is desirable that crates be made for shipment of sample containers. Full samples should be shipped express wherever possible and the agent should be advised that a red label shipment is being made consisting of oil or gas samples under pressure.

In order to obtain sufficient sample of gases at or near atmospheric pressure 18-20 liter heavy glass bottles sometimes are used. These are not recommended at a pressure less than about 2 cm. [5 in.] Hg vacuum due to the water leg required to balance atmospheric pressure.

SAMPLE CONTAINER CLEANING

Prior to taking samples, all sample containers should be properly cleaned. They should be placed in a vertical position with the bottom valve opened to drain all free liquid. They then should be thoroughly steamed out or washed with distillate, gasoline or cleaning solvent. If a solvent is used, the washing should continue until no color is shown on draining. Following a solvent rinse, fresh, clean water should be flowed upward through the cylinder and allowed to overflow out the top until the fluorescent oil film disappears.

If steam is used, the cleaning should continue until the effluent from the bottom is clear and without fluorescence. The cleaning of the container is an integral part of sampling and should always be done personally by the sampler wherever possible.

The practice of heating and then evacuating the container to clean it is not satisfactory. Tests have shown that all heavier ends may not be vaporized efficiently.

MECHANICS OF SAMPLING

The following summarizes suitable procedures for displacement sampling. Comparable procedures would apply for vacuum sampling.

Gas

If the sample is being taken from an orifice-meter manifold, bypass the meter and connect a length of tubing to the bleeder valve. Connect the other end to a tubing connection made up in the top valve of the sample container, but leave only hand tight for the moment. The container should be placed in a vertical position and the following procedure used:

1. Open the bleeder valve slightly, allowing enough gas to flow through the tubing to purge the air it contains.
2. Tighten the tubing nut with a wrench and allow the pressure in the tubing to build up to line pressure.
3. Open the bleeder valve and the top valve on the container completely and tighten the packing nuts if gas leaks are observed.
4. Crack the bottom valve on the sample container and allow the water to bleed off *Slowly.* Take at least 5 minutes to discharge the water.
5. After the water is discharged, let gas continue to flow through the bottom valve at the same rate for at least 15 to 20 minutes more.
6. *Do Not Let Gas Expand into the Bomb through the Top Valve. All Expansion Must Be to Atmosphere Via the Bottom Valve.*
7. Close the bottom valve, close top valve, and close bleeder valve in the precise order named. Disconnect tubing, dope pipe plugs with soap, and make up tight in both container valves. Tight valve packing nuts!
8. Head for the nearest water tank, slush pit, or puddle; immerse the whole container and check *Carefully* for leaks. Take your time with this check and look carefully for minute bubbles.

If the gas pressure is low and an 18-20 liter water bottle is to be used to take an atmospheric-pressure sample, the following procedure applies. Equipment needed includes a rubber stopper, or a wax-impregnated cork, to fit the bottle; a fairly long piece of tubing; about a 9-10 liter open container and a supply of clear, fresh water. Proceed as follows:

1. Fill the open container full of water and set it adjacent to the sample point. Fill the bottle with water, place palm of hand over opening and invert the bottle, immersing neck below

water level in open container. *If Air Bubbles Appear, Try it Again so there is no Air Trapped in the Bottle.*

2. Connect tubing to sample valve, crack valve to purge air from the tubing, then pass the free end of the tubing under the water level and up inside the bottle. Gas will then displace the water, which will overflow the bucket or tub.
3. Note the time required to displace the water from the bottle, and allow gas to pass into the bottle for a period four times as long. This is necessary to minimize the effects of absorption of carbon dioxide and hydrogen sulfide by the water.
4. Remove tubing and insert stopper in the bottle while the neck is still immersed. Make sure stopper is tight, then remove bottle from tub.
5. Wire stopper securely to bottle neck and, if possible, dry stopper and glass neck surfaces and coat with wax. Place in crate for shipment.

Liquid

Liquid samples should be taken from the liquid-level gauge glass according to the following procedure:

1. Close gauge-glass cocks, bleed off pressure, replace drain cock with suitable nipple or swage and a 1/4 in. barstock valve, if necessary. Close drain valve, open both cocks, and bleed liquid from the glass slowly via the drain valve to purge.
2. Connect copper tubing to drain valve and make up other end of tubing hand tight into top valve of sample container. Crack drain valve and allow liquid to purge air from tubing.
3. Gradually draw up tubing nut until it is tight. Turn drain valve and top valve on container wide open.
4. *Close the Top Gauge Cock.*
5. Crack bottom container valve and withdraw the equivalent of a full pint of water. Do this slowly, making sure a liquid level is always in sight in the gauge glass.
6. Close the bottom container valve. *Close the Bottom Gauge Cock. Open the Top Gauge Cock.*
7. Crack bottom container valve and *Slowly* withdraw the reminder of the water.
8. Close the bottom, top valves of the container, and the drain valve on the gauge glass in the order named. Disconnect tubing. Dope pipe plug threads with soap and pull up tight in valves. *Tighten Valve-Packing Nuts.*
9. Test carefully for leaks, under water, as discussed before.
10. Return gauge glass on separator to normal service.

The above procedure is preferred above all others since it provides sufficient sample at system pressure and temperature for accurate analysis, vastly simplifies laboratory procedure, and ensures safety in shipment by providing a substantial gas cap. Be sure to draw water before taking the sample and follow this procedure carefully.

With special reference to Steps 6 and 7 in the above procedure, make sure that the liquid level disappears when the bottom gauge cock is closed and the top cock opened. Failure to dissipate the liquid level means the bottom cock leaks and the container will be filled completely with liquid. This is wholly undesirable, so shut the bottom cock *Tight!*

When possible, the liquid and gas samples should be withdrawn simultaneously.

The equivalent procedures apply for vacuum sampling. In this case, of course, there is no displacement. However, once again it is important that a liquid sample container contain some gas cap above the liquid. After taking a liquid sample some hold the container vertically and crack the bottom valve so that a very small amount of liquid is drained out to ensure some vapor space.

Gas-Oil Ratio

When taking gas and liquid equilibrium samples from a separator, it is also necessary that the gas-oil ratio be obtained. Knowing this, the separator pressure, and temperature, these analyses may be recombined to yield a composite well stream.

In the ideal case where the residue gas is being sold, the gas-flow rate is simply obtained from the meter-run data. In the absence of the meter run, gas volumes may be determined using a critical-flow prover, orifice well tester, or pitot tube, using the test connection shown in Figure A.2.

Whenever the separator gauge pressure is above 100 kPa, the critical-flow prover is preferred because of its greater accuracy. The orifice well tester is relatively inaccurate, as is the pitot tube.

Except in those instances where a metering separator is available, the liquid flow rate from a vertical separator is most easily determined by measuring the rate of liquid rise in the separator. If valve D has been included (Figure A.2), this is easily done by:

1. Making two marks on the gauge glass at least several inches (10-20 cm) apart.
2. Manually reducing the liquid level below the lower mark.
3. Closing valve D and/or closing the liquid-level control valve.
4. Starting a stop watch when the liquid level reaches the lower mark.
5. Recording the time necessary for the liquid to reach the second mark and the distance between the marks.

The time elapsed should be several minutes for good accuracy. The distance of rise may be converted into volume of liquid by consulting a sales drawing of the separator manufacturer. If any appreciable quantity of water is being produced with the oil, the volume obtained above should be corrected. The rate of water production may be easily determined by using drain F and a container of known volume.

Whenever possible the incremental length used for the liquid determination should fall below the float flange (if any) in order to eliminate the error it might introduce by changing the separator's cross-sectional area.

In a horizontal separator the variation in volume with separator height is not linear, requiring a separate calculation. Once again after reference to the manufacturer's drawings, a calibration curve may be calculated for a given separator.

If not enough information is available for the calculation of volume, the only alternative is to shut down the separator, depressure, drain off oil, and calibrate using fresh water. It is actually a good practice to calibrate a separator in this manner at the time of installation so that the data are available. Once the samples are obtained they should be tagged, preferably with linen or other robust tags, and the following information included:

- Description of sample
- Location of separator
- Date, time
- Separator pressure
- Separator temperature
- Atmospheric temperature

If all of these standards are adhered to, the representative samples obtained will help immeasurably in the choice of proper operating conditions.

SPECIAL CONSIDERATIONS

The foregoing techniques are very suitable for most routine sampling problems. But, there are special cases where additional judgment is required.

Variations in Composition

Many flowing streams vary in rate and composition in a cyclical manner. One then needs an "average sampling." There are two practical choices: (1) take a lot of samples over a several day period, or (2) partially fill a single sample container with a series of withdrawals over a period of time. Either will work but the latter is less expensive.

Oftentimes the desired frequency of sampling can be established by observation. A common problem is surges of liquid entering the separation vessel. Notice the liquid level control pattern necessary to alleviate such surges. Set up a sampling frequency to "capture" an average sample.

Cold and Hot Samples

The sample container and its contents will change to ambient conditions. In this process part of a hot gas sample may condense or part of a cold liquid sample may vaporize. Sample conditions must be shown accurately so the laboratory can recreate them before withdrawing a sample for analysis.

Caution: Very high pressures can develop in a container full of liquid as it becomes hotter. In such case if the ambient is much above the sampling temperature bleed off a small amount of liquid to create a "gas cap" as noted earlier.

Storage Tanks

It is difficult to obtain a good sample from a stock tank near atmospheric pressure. First of all, the composition varies throughout the tank, particularly if the tank is rather full. The liquid head causes a relatively large pressure change from top to bottom. If tank vapor losses are negligible the best sample can be taken from the inlet line.

The gas analysis from a crude oil stock tank may vary widely over a 24 hour period due to "breathing" characteristics and weathering.

APPENDIX B
General Conversion Factors

Base SI metric units

Quantity	Name	Symbol
Length	meter	m
Mass	kilogram	kg
Time	second	s
Electric current	ampere	A
Thermodynamic temperature	kelvin	K
Luminous intensity	candela	cd
Amount of substance	mole	mol

Important derived and supplementary SI units

Quantity	Name	Symbol	Formula
Electric capacitance	farad	F	A-s/V
Electric charge	coulomb	C	A-s
Electric conductance	siemens	S	A/V
Electric inductance	henry	H	Wb/A
Electric potential	volt	V	W/A
Electric quantity	coulomb	C	A-s
Electric resistance	ohm	Ω	V/A
Electromotive force	volt	V	W/A
Energy	joule	J	N-m
Energy flux	watt	W	J/s
Force	newton	N	kg-m/s^{2}
Frequency	hertz	Hz	1/s
Illuminance	lux	1x	1m/m^{2}
Luminous flux	lumen	1m	cd-sr
Magnetic flux	weber	Wb	V-s
Magnetic flux density	tesla	T	Wb/m^{2}
Power	watt	W	J/s
Pressure	pascal	Pa	N/m^{2}
Quantity of heat	joule	J	N-m
Solid Angle	steradian	sr	—
Work	joule	J	N-m

SI prefixes and multiplication factors

Prefix	SI Symbol	Factor
tera	T	10^{12}
giga	G	10^{9}
mega	M	10^{6}
kilo	k	10^{3}
hecto	h	10^{2}
deka	da	10^{1}
deci	d	10^{-1}
centi	c	10^{-2}
milli	m	10^{-3}
micro	μ	10^{-6}
nano	n	10^{-9}
pico	p	10^{-12}
femto	f	10^{-15}
atto	a	10^{-18}

To convert from	to	Multiply by
acre	hectare	0.40469
acre	meter^{2}	4,046.9
acre-feet	meter^{3}	1,233.5
ampere-hour	coulomb	3,600.0
angstrom	nanometer	0.10000
astronomical unit	gigameter	149.598
atmosphere	bar	1.01325
atmosphere (normal)	kilopascal	101.325
atmosphere(normal)	pascal	101,325
atmosphere (1 kgf/cm^{3})	pascal	98,067
bar	kilopascal	100.000
bar	pascal	100,000
barrel (42 US gal)	meter^{3}	0.15899
barrel/acre-foot	decimeter^{3}/meter^{3}	0.12889
barrel/day	decimeter^{3}/second	0.00184
barrel/day	meter^{3}/day	0.15899
barrel/day (1 kg/dm^{3})	tonne/annum	58.0304
barrel/foot	meter^{3}/meter	0.52161
barrel/hour	decimeter^{3}/second	0.04416
barrel/hour	meter^{3}/hour	0.15899
barrel/US ton	meter^{3}/tonne	0.17525
board-foot	meter^{3}	0.00236
British thermal unit	joule	1,055.1
British thermal unit	kilojoule	1.05506
British thermal unit	watt-hour	0.29288
BTU/bhp-hour	watt/kilowatt	0.39301
BTU/foot^{3}	kilojoule/meter^{3}	37.2590
BTU/foot^{3}	kilowatt-hour/meter^{3}	0.01035
BTU/US gallon	kilojoule/meter^{3}	278.716
BTU/US gallon	kilowatt-hour/meter^{3}	0.07742
BTU/hour	watt	0.29307
BTU/minute	watt	17.5843
BTU/second	kilowatt	1.05506
BTU/hour-foot^{2}	watt/meter^{2}	3.15459
BTU/hour-foot^{3}	watt/meter^{3}	10.3497
BTU/hr-foot^{2}-°F	watt/meter^{2}-kelvin	5.67826
BTU/second-foot^{3}-°F	kilowatt/meter^{2}-kelvin	20.4418
BTU/hour-foot^{3}-°F	watt/meter^{3}-kelvin	18.6295
BTU/second-foot^{3}-°F	kilowatt/meter^{3}-kelvin	67.0661
BTU/hour-foot^{2}-°F/foot	watt/meter-kelvin	1.73074
BTU/hour-foot^{2}-°F/inch	watt/meter-kelvin	0.14428
BTU/second-foot^{2}-°F/in.	watt/meter-kelvin	519.220
BTU/pound mass	joules/gram	2.32600
BTU/pound mass	watt-hour/kilogram	0.64611
BTU/pound mol	joules/mol	2.32600
BTU/pound-°F	joules/gram-kelvin	4.18680
BTU/pound mol-°F	joules/mol-kelvin	4.18680
bushel	decimeter^{3}	35.2391
bushel	meter^{3}	0.03524
calorie	joule	4.18400
calorie/pound	joule/kilogram	9.22414
centipoise	newton-second/meter^{2}	0.00100
centipoise	pascal-second	0.00100
centistroke	millimeter^{2}/second	1.00000
chain	meter	20.1168
cycle/second	hertz	1.00000
darcy	micrometer^{2}	0.98692
degree (angle)	radian	0.01745
°F/100 foot	millikelvin/meter	18.2269
°F.-foot^{2}-hour/BTU	kelvin-meter^{2}/watt	0.17611
dyne	millinewton	0.01000
erg	microjoule	0.10000
erg	millijoule	0.00010
fathom	meter	1.82880
foot	centimeter	30.4800
foot	meter	0.30480
foot/day	meter/day	0.30480
foot/°F.	meter/kelvin	0.54864
foot/hour	meter/hour	0.30480
foot/hour	millimeter/second	0.08467
foot/minute	centimeter/second	0.50800
foot/minute	meter/minute	0.30480
foot/minute	meter/second	0.00508
foot/second	meter/second	0.30480
foot/second^{2}	meter/second^{2}	0.30480
foot^{2}	centimeter^{2}	929.030
foot^{2}	meter^{2}	0.09290
foot^{2}/hour	millimeter^{2}/second	25.8064
foot^{2}/second	millimeter^{2}/second	92.903
foot^{3}	decimeter^{3}	28.3169
foot^{3}	meter^{3}	0.02832
foot^{3}/day	decimeter^{3}/second	0.00033
foot^{3}/day	meter^{3}/day	0.02832
foot^{3}/day	meter^{3}/hour	0.00118

To convert from	to	Multiply by
foot^3/foot	meter^3/meter	0.09290
foot^3/hour	decimeter^3/second	0.00787
foot^3/hour	meter^3/hour	0.02832
foot^3/minute	decimeter^3/second	0.47195
foot^3/minute	meter^3/minute	0.02832
foot^3/pound	meter^3/kilogram	0.06243
foot^{3}pound mol	meter^3/kilomol	0.06243
foot^3/second	meter^3/second	0.02832
footcandle	lux	10.0764
footcandle-second	lux-second	10.0764
foot-pound force	joule	1.35582
foot-pound/US gallon	kilojoule/meter^3	0.35817
foot-pound/minute	milliwatt	22.5970
foot-pound/minute	watt	0.02260
foot-pound/second	watt	1.35582
gallon US liquid	meter^3	0.00379
gallon US/foot	meter^3/meter	0.01242
gallon US/horsepower-hr.	decimeter^3/megajoule	1.41009
gallon US/horsepower-hr.	millimeter^3/joule	1.41009
gallon US/1,000 barrel	centimeter^3/meter^3	23.8095
gallon US/foot^3	decimeter^3/meter^3	133.681
gallon US/hour	decimeter^3/second	0.00105
gallon US/hour	meter^3/hour	0.00379
gallon US/mile	decimeter^3/100 km	235.215
gallon US/minute	decimeter^3/second	0.06309
gallon US/minute	meter^3/hour	0.22712
gallon US/ton US	decimeter^3/tonne	4.17270
grain	milligram	64.7989
grain/gallon US	gram/meter^3	17.1181
grain/100 foot^3	milligram/meter^3	22.8835
horsepower	kilowatt	0.74570
horsepower (boiler)	kilowatt	9.80950
horsepower/foot^3	kilowatt/meter^3	26.3341
horsepower-hour	kilowatt/hour	0.74570
horsepower-hour	megajoule	2.68452
inch	centimeter	2.54000
inch	millimeter	25.4000
inch	pica	6.02250
inch	point	72.2700
inch Hg (60°F)	kilopascal	3.37685
inch H$_2$0 (60°F)	kilopascal	0.24884
inch/minute	centimeter/minute	2.54000
inch/second	centimeter/second	2.54000
inch/second	millimeter/second	25.4000
inch/year	millimeter/annum	25.4000
inch^2	centimeter^2	6.45160
inch^2	millimeter^2	645.160
inch^2/second	millimeter^2/second	645.160
inch^3	centimeter^3	16.3871
inch^4	centimeter^4	41.6231
joule	British thermal unit	0.00095
joule	foot-pound force	0.73756
kilogram	pound mass av.	2.20462
kilogram/meter^3	pound mass/foot^3	0.06243
kilometer/hour	mile/hour statute	0.62137
kilowatt-hour	kilojoule	3,600.0
kilowatt-hour	megajoule	3.60000
knot	kilometer/hour	1.85200
lambert	candela/meter^2	3,183.1
link	meter	0.20117
meter	foot	3.28084
meter^2	foot^2	10.7639
meter^3	barrel (42 US gal)	6.28981
meter^3	foot^3	35.3147
meter/minute	foot/minute	3.28084
meter^3/minute	gallon US/minute	264.172
micron	micrometer	1.00000
mil	micrometer	25.4000
mile,nautical	kilometer	1.85200
mile US statute	kilometer	1.60934
mile^2 US statute	kilometer^2	2.58999
mile (US stat.)/US gal	kilometer/decimeter^3	0.42514
mile (US stat.)/hour	kilometer/hour	1.60934
minute (angle)	radian	0.00029
newton	pound force av.	0.22481
ounce force av	newton	0.27801
ounce mass av	gram	28.3495
ounce mass av	kilogram	0.02835
ounce US fluid	centimeter^3	29.5735
ounce troy	gram	31.1035
part/million (volume)	centimeter^3/meter^3	1.00000
pascal	pound force/foot^2	0.02089
pascal	pound force/inch^2	0.00015

To convert from	to	Multiply by
pica	centimeter	0.42175
pica	inch	0.16604
pint US dry	decimeter^3	0.55061
pint US liquid	decimeter^3	0.47318
point	centimeter	0.03515
point	inch	0.01384
pound force	newton	4.44822
pound mass av	kilogram	0.45359
pound mass troy	kilogram	0.37324
pound mol	kilomol	0.45359
pound force/foot^2	pascal	47.8803
pound force/inch^2	kilopascal	6.89476
pound force/inch^2/foot	kilopascal/meter	22.6206
pound mass/barrel	gram/decimeter^3	2.85301
pound mass/barrel	kilogram/meter^3	2.85301
pound mass/foot	kilogram/meter	1.48816
pound mass/foot^2	kilogram/meter^2	4.88243
pound mass/foot^3	kilogram/meter^3	16.0185
pound mass/US gallon	gram/centimeter^3	0.11983
pound mass/US gallon	kilogram-decimeter^3	0.11983
pound mass/horsepower-hr.	kilogram/kilowatt-hour	0.60828
pound mass/horsepower-hr.	kilogram/megajoule	0.16897
pound mass/horsepower-hr.	milligram/joule	0.16897
pound mass/hour	tonne/day	0.45359
pound mass/minute	kilogram/minute	0.45359
pound mass/second	kilogram/second	0.45359
pound mass/hour-foot	gram/second-meter	0.41338
pound mass/second-foot	kilogram/second-meter	1.48816
pound mass/second-foot	pascal-second	1.48816
pound mass/second foot^2	kilogram/second-meter^2	4.88243
pound mol	kilomol	0.45359
pound mol/foot^3	kilomol/meter^3	16.0185
pound mol/US gallon	kilomol/meter^3	119.826
pound mol/hour	kilomol/hour	0.45359
pound mol/second	kilomol/second	0.45359
pound force-foot	joule	1.35582
pound force-foot	newton-meter	1.35582
pound force-inch	joule	0.11298
pound force-inch	newton-meter	0.11298
pound force-foot^2	kilogram-meter^2	0.04214
pound force-foot/inch	newton-meter/meter	53.3787
pound force-foot/inch	newton-meter/meter	4.44822
pound force-foot/inch^2	joule/centimeter	0.00210
pound force-foot/second	kilogram-meter/second	0.13826
pound force-second/foot^2	pascal-second	47.8803
quart US	decimeter^3	0.94635
radian (angle)	degree	57.2958
radian (angle)	revolution	0.15915
radian (angle)/minute	revolution/minute	0.15915
°Rankin	kelvin	0.55556
revolution	radian (angle)	6.28319
revolution/minute	radian/minute	6.28319
section	hectare	258.999
therm	megajoule	105.506
ton force US	kilonewton	8.89644
ton mass US short	kilogram	907.185
ton mass US short	toone	0.90718
tonne	kilogram	1,000.0
tonne	megagram	1.00000
ton force US/foot^2	megapascal	0.09576
ton force US/inch^2	megapascal	13.7895
ton force US-mile	megajoule	14.3174
ton mass US/day	tonne/hour	0.03780
ton mass US/hour	tonne/day	0.90718
ton mass US/day	kilogram/second	0.01050
ton mass US/hour	kilogram/second	0.25120
ton mass US/minute	kilogram/second	15.1197
ton mass US/year	tonne-annum	0.90718
ton mass US/foot^2	tonne/meter^2	9.76486
torr	pascal	133.322
watt	BTU/minute	0.05687
watt	joule/second	1.00000
watt-hour	joule	3,600.0
watt-second	joule	1.00000
yard	meter	0.91440
yard^2	meter^2	0.83613
yard^3	meter^3	0.76455
year	annum	1.00000

This	times this	gives this
Acre	43,560	sq. ft.
Acre-feet	7,757.8	bbl.(42 gal.)
"	43,560	cu. ft.
Ampere-hours	3,600	Coulombs
Angstroms	0.0001	Microns
Atm. at 32°F.)	33.90	ft. of water
"	1,034	g./sq.cm.
"	29.92	in. Hg
"	760.18	mm. Hg
"	14.697	psi.
Bbl.(42 gal.)	5.6146	cu. ft.
"	9,702.03	cu. in.
Bbl./day(42 gal.)	1.75	g.p.h.
"	0.0292	g.p.m.
Bbl./hr.(42 gal.)	0.0936	cu. ft./min.
"	0.7	g.p.m.
Board feet	144.0	cu. in.
B.t.u.	778	ft.-lb.
"	252	gram-cal.
"	0.0003927	hp.-hr.
"	0.0002928	kw.-hr.
B.t.u./min.	46,681.68	ft.-lb./hr.
"	778	ft.-lb./min.
"	12.967	ft.-lb./sec.
"	0.02356	hp.
"	0.01758	kw.
B.t.u./sec.	46,681.68	ft.-lb./min.
"	778	ft.-lb./sec.
"	1.4146	hp.
"	1.05487	kw.
Bushels(U.S. std.)	1.2445	cu. ft.
"	2,150	cu. in.
Candles/sq. in.	452.39	ft.-Lamberts
"	0.48695	Lamberts
Candlepower (sph.)	12.566	lumens
Carat (intl.)	3.08647	grains
"	0.2	g.
"	200	mg.
Centimeters	0.03281	ft.
"	0.3937	in.
"	393.7	mils
Cm. of mercury	0.01316	atm.
"	0.4461	ft. of water
"	27.85	lb./sq. ft.
"	0.1934	psi.
Cm. per second	1.9685	ft./min.
"	0.03281	ft./sec.
"	0.02237	m.p.h.
Centipoise	1.0000	g./cm.-sec.
"	0.00067197	lb./ft.-sec.
Circular mils	0.7854	sq. mils
Coulombs	0.1	Abcoulombs
"	0.0002778	amp.-hr.
Cubic centimeters	0.00003531	cu. ft.
"	0.0002642	gal. (U.S. liq.)
"	0.001	liters
Cubic feet	0.00002296	acre-ft.
"	0.1781	bbl. (42 gal.)
"	0.80356	bu.
"	28,320	cc.
"	1,728	cu. in.
"	7,481	gal. (U.S. liq.)
"	28.32	liters
Cu. ft., million	22.95	acre-ft.
Cu. ft. per min	10.686	bbl.(42 gal.)/hr.
"	0.1781	bbl.(42 gal.)/min.
"	1,440	cu. ft./day
"	7,481	g.p.m. (U.S. liq.)
Cubic inches	0.000465	bu.
"	16.39	cc.
"	0.0005787	cu. ft.
"	0.004329	gal.(U.S. liq.)
"	0.016387	liters
Degrees of arc	60	minutes
"	0.017453	radians
Dynes	0.0010197	g.
"	0.000002248	lb.
Dyne-cm.	1	ergs
"	0.00000007376	ft.-lb.
"	0.0010197	g.-cm.
Ems	0.1666	in.
Ergs	1	dyne-cm.
Faradays	26.80	amp.-hr.
Fathoms	6	ft.
Feet	30.48	cm.
"	0.16667	fathoms
"	0.000164468	mile(U.S. naut.)
"	0.000189394	mile(U.S. stat.)
Feet per minute	0.5080	cm./sec.
"	0.00987	knots/hr.
"	0.01136	m.p.h.
Feet per second	0.68121	m.p.h.
"	0.01136	miles/min.
Feet of water	0.02950	atm.
"	0.4335	psi.
Foot-candles	1	lumens/sq. ft.
Foot-Lamberts	0.0022105	candles/sq. in.
"	1	lumens/sq. ft.
Foot-pounds	0.0012853	B.t.u.
"	0.000000505	hp.-hr.
"	1.35582	joules
"	0.0003241	kg.-cal.
"	0.0000003766	kw.-hr.
Foot-pounds/hour	0.001284	B.t.u./hr.
"	0.00002141	B.t.u./min
"	0.000000357	B.t.u./sec
"	0.01666	ft.-lb./min.
"	0.000000505	hp.
"	0.0000003766	kw.
Foot-pounds/min.	0.077118	B.t.u./hr.
"	0.001286	B.t.u./min.
"	0.0000303	hp.
"	0.00002256	kw.
Foot-pounds/sec.	0.001285	B.t.u./sec.
"	0.001818	hp.
Furlong	660	ft.

This	times this	gives this
Gal. (U.S. dry)	0.15556	cu. ft.
"	1.1637	gal. (U.S. liq.)
Gal. (U.S. liq.)	0.02381	bbl. (42 gal.)
"	0.133681	cu. ft.
"	231	cu. in.
"	0.859365	gal. (U.S. dry)
"	3.7854	liters
Gal./hr. (U.S. liq.)	0.1337	cu. t./hr.
"	0.002228	cu. ft./min.
"	0.01666	g.p.m. (U.S. liq.)
G.p.m. (U.S. liq.)	34.2857	bbl./day (42 gal.)
"	1.4286	bbl./hr. (42 gal.)
"	0.02381	bbl./min. (42 gal.)
"	192.49	cu. ft./day
"	0.1337	cu. ft./min.
"	1,440	gal./day
Grains	0.00014286	lb.,avoir.
Grains/gal. (U.S. liq.)	0.01714	g./l./
"	17.118	p.p.m.
Grams	980.7	dynes
"	0.035274	oz.,avoir.
"	0.0022046	lb.,avoir.
Gram-calories	0.003968	B.t.u.
"	3.0875	ft.-lb.
Grams per liter	58.418	grains/gal
"	1,000	p.p.m.
"	0.0624	lb./cu. ft.
Horsepower	2,544	B.t.u./hr.
"	42.44	B.t.u./min.
"	33,000	ft.-lb./min.
"	550	ft.-lb./sec.
"	745.70	joules/sec.
"	0.74570	kw.
Horsepower-hours	2,544	B.t.u.
"	1,980,000	ft.-lb.
"	0.7457	kw.-hr.
Inches	2.54	cm.
"	0.08333	ft.
"	25,400.05	microns
"	1,000	mils
Inches of mercury	0.03242	atm.
"	0.49116	psi.
Joules	0.0009486	B.t.u.
"	0.239	calories
"	0.73756	ft.-lb.
Kilograms	2.205	lb.,avoir.
Kilogram-calories	3.9683	B.t.u.
"	3,087.5	ft.-lb.
Kilowatts	56.92	B.t.u./min.
"	0.948	B.t.u./sec.
"	44,253	ft.-lb./min.
"	1.341	hp.
Kilowatt-hours	3,415	B.t.u.
"	2,655,199	ft.-lb.
"	1.341	hp.-hr.
Knots (naut.mile)	6,080	ft.
"	1.152	mile (U.S. stat.)
Knots	101.34	ft./min.
"	1.152	m.p.h. (U.S. stat.)
Liters	1,000	cc.
"	0.03531	cu. ft.
"	61.0271	cu. in.
"	0.26418	gal. (U.S. liq.)
"	1.057	qt. (U.S. liq.)
Logarithm, com	2.3026	logarithm,Nap.
Logarithm, Nap	0.4343	logarithm, com
Meters	3.2808	ft.
"	39.37	in.
"	1,000,000	microns
Microns	10,000	angstroms
"	0.00003937	in.
"	0.001	mm.
Mils	0.00254	cm.
"	0.001	in.
"	25.4	microns
Miles (U.S. stat.)	5,280	ft.
"	1.6093	km.
Miles (U.S. naut.)	6,080	ft.
Miles per hour	44.704	cm./sec.
"	88	ft./min.
"	1.4667	ft./sec.
"	0.01666	miles/min.
Milliliters	0.061	cu. in.
Millimeters	0.03937	in.
"	39.37	mils
Minutes of arc	0.016667	deg. of arc
Ounces (avoir.)	28.35	g.
Parts per million	0.05835	grains/gal.
"	0.001	g./l.
Points	0.013889	in.
Pounds, avoir.	444,823	dynes
"	7,000	grains
"	453.59	g.
"	1.215	lb., Troy, Apoth
Pounds per gallon	7.4805	lb./cu. ft.
Pounds per cu. ft.	0.016018	g./cc.
"	0.13368	lb./gal. (U.S. liq.)
Pounds per sq. in.	0.06804	atm.
"	5.1715	cm. Hg.
"	2.307	ft. of water
"	2.036	in. Hg.
"	144	lb./sq. ft.
Quarts	0.946	liters
Radians	57.2958	deg. of arc
Radians per sec.	9.549	r.p.m.
Rods	16.5	ft.
"	5.5	yd.
Sections	640	acres
"	1	sq. miles
Square feet	0.00002296	acres
"	144	sq. in.
Square mils	1.2732	circular mils
Township	23,040	acres
"	36	sections
"	36	sq. miles
Vara	2.7777	ft.

°C	Temp. °C or °F -459.69 to -4	°F
-273.16	-459.69	
-267.78	-450	
-262.22	-440	
-256.67	-430	
-251.11	-420	
-245.56	-410	
-240.00	-400	
-234.44	-390	
-228.89	-380	
-223.33	-370	
-217.78	-360	
-212.22	-350	
-206.67	-340	
-201.11	-330	
-195.56	-320	
-190.00	-310	
-184.44	-300	
-178.89	-290	
-173.33	-280	
-169.53	-273.16	-459.7
-168.89	-272	-457.6
-167.78	-270	-454.0
-162.22	-260	-436.0
-156.67	-250	-418.0
-151.11	-240	-400.0
-145.56	-230	-382.0
-140.00	-220	-364.0
-134.44	-210	-346.0
-128.89	-200	-328.0
-123.33	-190	-310.0
-117.78	-180	-292.0
-112.22	-170	-274.0
-106.67	-160	-256.0
-101.11	-150	-238.0
-95.56	-140	-220.0
-90.00	-130	-202.0
-84.44	-120	-184.0
-78.89	-110	-166.0
-73.33	-100	-148.0
-70.56	-95	-139.0
-67.78	-90	-130.0
-65.00	-85	-121.0
-62.22	-80	-112.0
-59.45	-75	-103.0
-56.67	-70	-94.0
-53.89	-65	-85.0
-51.11	-60	-76.0
-48.34	-55	-67.0
-45.56	-50	-58.0
-42.78	-45	-49.0
-40.00	-40	-40.0
-39.45	-39	-38.2
-38.89	-38	-36.4
-38.34	-37	-34.6
-37.78	-36	-32.8
-37.23	-35	-31.0
-36.67	-34	-29.2
-36.12	-33	-27.4
-35.56	-32	-25.6
-35.00	-31	-23.8
-34.44	-30	-22.0
-33.89	-29	-20.2
-33.33	-28	-18.4
-32.78	-27	-16.6
-32.22	-26	-14.8
-31.67	-25	-13.0
-31.11	-24	-11.2
-30.56	-23	-9.4
-30.00	-22	-7.6
-29.45	-21	-5.8
-28.89	-20	-4.0
-28.34	-19	-2.2
-27.78	-18	-0.4
-27.23	-17	1.4
-26.67	-16	3.2
-26.12	-15	5.0
-25.56	-14	6.8
-25.00	-13	8.6
-24.44	-12	10.4
-23.89	-11	12.2
-23.33	-10	14.0
-22.78	-9	15.8
-22.22	-8	17.6
-21.67	-7	19.4
-21.11	-6	21.2
-20.56	-5	23.0
-20.00	-4	24.8

°C	Temp. °C or °F -3 to 83	°F
-19.44	-3	26.6
-18.89	-2	28.4
-18.33	-1	30.2
-17.8	0	32.0
-17.2	1	33.8
-16.7	2	35.6
-16.1	3	37.4
-15.6	4	39.2
-15.0	5	41.0
-14.4	6	42.8
-13.9	7	44.6
-13.3	8	46.4
-12.8	9	48.2
-12.2	10	50.0
-11.7	11	51.8
-11.1	12	53.6
-10.6	13	55.4
-10.0	14	57.2
-9.44	15	59.0
-8.89	16	60.8
-8.33	17	62.6
-7.78	18	64.4
-7.22	19	66.2
-6.67	20	68.0
-6.11	21	69.8
-5.56	22	71.6
-5.00	23	73.4
-4.44	24	75.2
-3.89	25	77.0
-3.33	26	78.8
-2.78	27	80.6
-2.22	28	82.4
-1.67	29	84.2
-1.11	30	86.0
-0.56	31	87.8
0	32	89.6
0.56	33	91.4
1.11	34	93.2
1.67	35	95.0
2.22	36	96.8
2.78	37	98.6
3.33	38	100.4
3.89	39	102.2
4.44	40	104.0
5.00	41	105.8
5.56	42	107.6
6.11	43	109.4
6.67	44	111.2
7.22	45	113.0
7.78	46	114.8
8.33	47	116.6
8.89	48	118.4
9.44	49	120.2
10.0	50	122.0
10.6	51	123.8
11.1	52	125.6
11.7	53	127.4
12.2	54	129.2
12.8	55	131.0
13.3	56	132.8
13.9	57	134.6
14.4	58	136.4
15.0	59	138.2
15.6	60	140.0
16.1	. 61	141.8
16.7	62	143.6
17.2	63	145.4
17.8	64	147.2
18.3	65	149.0
18.9	66	150.8
19.4	67	152.6
20.0	68	154.4
20.6	69	156.2
21.1	70	158.0
21.7	71	159.8
22.2	72	161.6
22.8	73	163.4
23.3	74	165.2
23.9	75	167.0
24.4	76	168.8
25.0	77	170.6
25.6	78	172.4
26.1	79	174.2
26.7	80	176.0
27.2	81	177.8
27.8	82	179.6
28.3	83	181.4

°C	Temp. °C or °F 84 to 800	°F
28.9	84	183.2
29.4	85	185.0
30.0	86	186.8
30.6	87	188.6
31.1	88	190.4
31.7	89	192.2
32.2	90	194.0
32.8	91	195.8
33.3	92	197.6
33.9	93	199.4
34.4	94	201.2
35.0	95	203.0
35.6	96	204.8
36.1	97	206.6
36.7	98	208.4
37.2	99	210.2
37.8	100	212.0
43.3	110	230.0
48.9	120	248.0
54.4	130	266.0
60.0	140	284.0
65.6	150	302.0
71.1	160	320.0
76.7	170	338.0
82.2	180	356.0
87.8	190	374.0
93.3	200	392.0
98.9	210	410.0
104.4	220	428.0
110.0	230	446.0
115.6	240	464.0
121.1	250	482.0
126.7	260	500.0
132.2	270	518.0
137.8	280	536.0
143.3	290	554.0
148.9	300	572.0
154.4	310	590.0
160.0	320	608.0
165.6	330	626.0
171.1	340	644.0
176.7	350	662.0
182.2	360	680.0
187.8	370	698.0
193.3	380	716.0
198.9	390	734.0
204.4	400	752.0
210.0	410	770.0
215.6	420	788.0
221.1	430	806.0
226.7	440	824.6
232.2	450	842.0
237.8	460	860.0
243.3	470	878.0
248.9	480	896.0
254.4	490	914.0
260.0	500	932.0
265.6	510	950.0
271.1	520	968.0
276.7	530	986.0
282.2	540	1,004.0
287.8	550	1,022.0
293.3	560	1,040.0
298.9	570	1,058.0
304.4	580	1,076.0
310.0	590	1,094.0
315.6	600	1,112.0
321.1	610	1,130.0
326.7	620	1,148.0
332.2	630	1,166.0
337.8	640	1,184.0
343.3	650	1,202.0
248.9	660	1,220.0
354.4	670	1,238.0
360.0	680	1,256.0
365.6	690	1,274.0
371.1	700	1,292.0
376.7	710	1,310.0
382.2	720	1,328.0
387.8	730	1,346.0
393.3	740	1,364.0
398.9	750	1,382.0
404.4	760	1,400.0
410.0	770	1,418.0
415.6	780	1,436.0
421.1	790	1,454.0
426.7	800	1,472.0

°C	Temp. °C or °F 810 to 1,670	°F
432.2	810	1,490.0
437.8	820	1,508.0
443.3	830	1,526.0
448.9	840	1,544.0
454.4	850	1,562.0
460.0	860	1,580.0
465.6	870	1,598.0
471.1	880	1,616.0
476.7	890	1,634.0
482.2	900	1,652.0
487.8	910	1,670.0
493.3	920	1,688.0
498.9	930	1,706.0
504.4	940	1,724.0
510.0	950	1,742.0
515.6	960	1,760.0
521.1	970	1,778.0
526.7	980	1,796.0
532.2	990	1,814.0
537.8	1,000	1,832.0
543.3	1,010	1,850.0
548.9	1,020	1,868.0
554.4	1,030	1,886.0
560.0	1,040	1,904.0
565.6	1,050	1,922.0
571.1	1,060	1,940.0
576.7	1,070	1,958.0
582.2	1,080	1,976.0
587.8	1,090	1,994.0
593.3	1,100	2,012.0
598.9	1,110	2,030.0
604.4	1,120	2,048.0
610.0	1,130	2,066.0
615.6	1,140	2,084.0
621.1	1,150	2,102.0
626.7	1,160	2,120.0
632.2	1,170	2,138.0
637.8	1,180	2,156.0
643.3	1,190	2,174.0
648.9	1,200	2,192.0
654.4	1,210	2,210.0
660.0	1,220	2,228.0
665.6	1,230	2,246.0
671.1	1,240	2,264.0
676.7	1,250	2,282.0
682.2	1,260	2,300.0
687.8	1,270	2,318.0
693.3	1,280	2,336.0
698.9	1,290	2,354.0
704.4	1,300	2,372.0
710.0	1,310	2,390.0
715.6	1,320	2,408.0
721.1	1,330	2,426.0
726.7	1,340	2,444.0
732.2	1,350	2,462.0
737.8	1,360	2,480.0
743.3	1,370	2,498.0
748.9	1,380	2,516.0
754.4	1,390	2,534.0
760.0	1,400	2,552.0
765.6	1,410	2,570.0
771.1	1,420	2,588.0
776.7	1,430	2,606.0
782.2	1,440	2,624.0
787.8	1,450	2,642.0
793.3	1,460	2,660.0
798.9	1,470	2,678.0
804.4	1,480	2,696.0
810.0	1,490	2,714.0
815.6	1,500	2,732.0
821.1	1,510	2,750.0
826.7	1,520	2,768.0
832.2	1,530	2,786.0
837.8	1,540	2,804.0
843.3	1,550	2,822.0
848.9	1,560	2,840.0
854.4	1,570	2,858.0
860.0	1,580	2,876.0
865.6	1,590	2,894.0
871.1	1,600	2,912.0
876.7	1,610	2,930.0
882.2	1,620	2,948.0
887.8	1,630	2,966.0
893.3	1,640	2,984.0
898.9	1,650	3,002.0
904.4	1,660	3,020.0
910.0	1,670	3,038.0

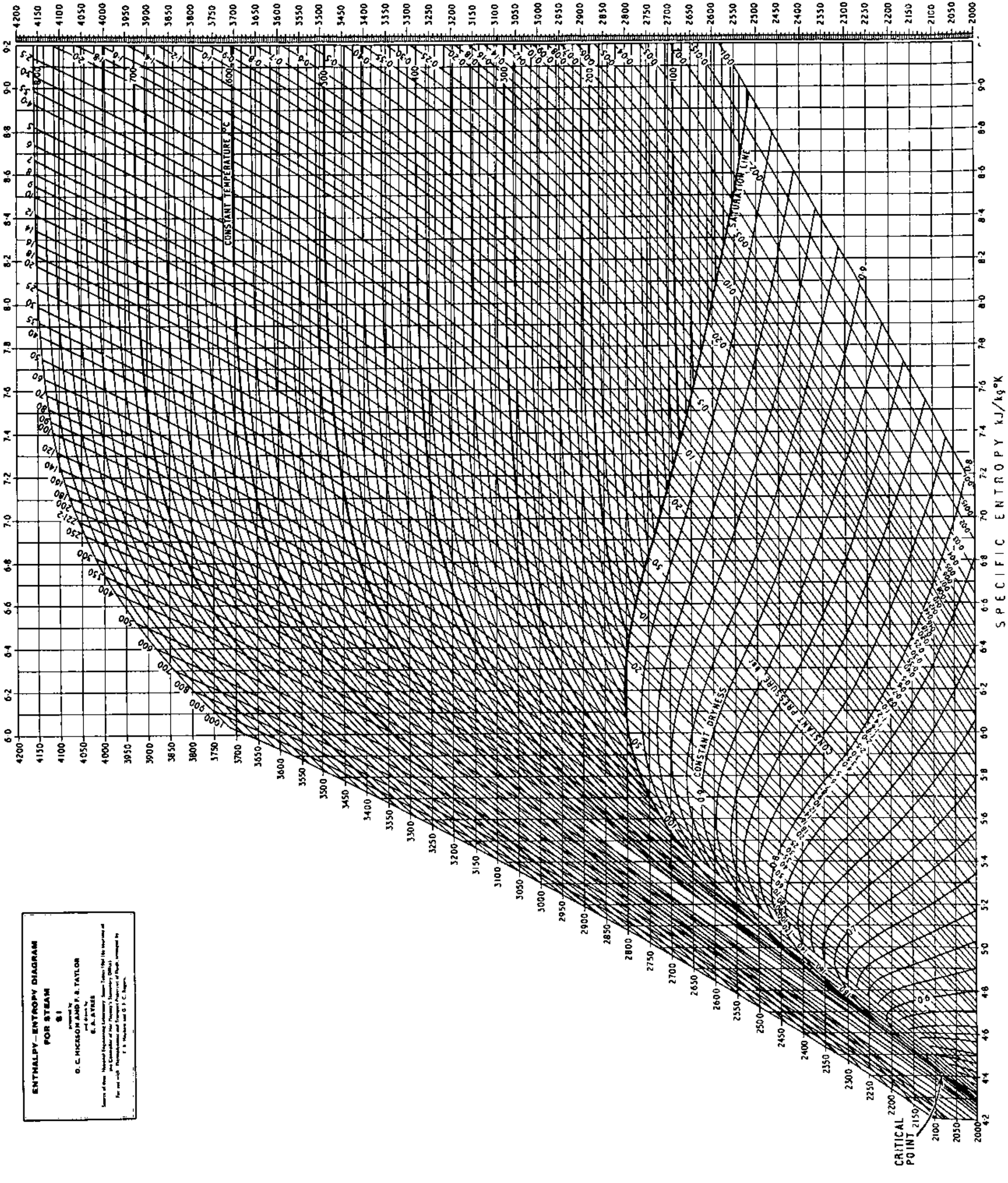
°C	Temp. °C or °F 1,680 to 3,000	°F
915.6	1,680	3,056.0
921.1	1,690	3,074.0
926.7	1,700	3,092.0
932.2	1,710	3,110.0
937.8	1,720	3,128.0
943.3	1,730	3,146.0
948.9	1,740	3,164.0
954.4	1,750	3,182.0
960.0	1,760	3,200.0
965.6	1,770	3,218.0
971.1	1,780	3,236.0
976.7	1,790	3,254.0
982.2	1,800	3,272.0
987.8	1,810	3,290.0
993.3	1,820	3,308.0
998.9	1,830	3,326.0
1,004.4	1,840	3,344.0
1,010.0	1,850	3,362.0
1,015.6	1,860	3,380.0
1,021.1	1,870	3,398.0
1,026.7	1,880	3,416.0
1,032.2	1,890	3,434.0
1,037.8	1,900	3,452.0
1,043.3	1,910	3,470.0
1,048.9	1,920	3,488.0
1,054.4	1,930	3,506.0
1,060.0	1,940	3,524.0
1,065.6	1,950	3,542.0
1,071.1	1,960	3,560.0
1,076.7	1,970	3,578.0
1,082.2	1,980	3,596.0
1,087.8	1,990	3,614.0
1,093.3	2,000	3,632.0
1,098.9	2,010	3,650.0
1,104.4	2,020	3,668.0
1,110.0	2,030	3,686.0
1,115.6	2,040	3,704.0
1,121.1	2,050	3,722.0
1,126.7	2,060	3,740.0
1,132.2	2,070	3,758.0
1,137.8	2,080	3,776.0
1,143.3	2,090	3,794.0
1,148.9	2,100	3,812.0
1,154.4	2,110	3,830.0
1,160.0	2,120	3,848.0
1,165.6	2,130	3,866.0
1,171.1	2,140	3,884.0
1,176.7	2,150	3,902.0
1,182.2	2,160	3,920.0
1,187.8	2,170	3,938.0
1,193.3	2,180	3,956.0
1,198.9	2,190	3,974.0
1,204.4	2,200	3,992.0
1,210.0	2,210	4,010.0
1,215.6	2,220	4,028.0
1,221.1	2,230	4,046.0
1,226.7	2,240	4,064.0
1,232.2	2,250	4,082.0
1,237.8	2,260	4,100.0
1,243.3	2,270	4,118.0
1,248.9	2,280	4,136.0
1,254.4	2,290	4,154.0
1,260.0	2,300	4,172.0
1,265.6	2,310	4,190.0
1,271.1	2,320	4,208.0
1,276.7	2,330	4,226.0
1,282.2	2,340	4,244.0
1,287.8	2,350	4,262.0
1,293.3	2,360	4,280.0
1,298.9	2,370	4,298.0
1,304.4	2,380	4,316.0
1,310.0	2,390	4,334.0
1,343.3	2,450	4,442.0
1,371.1	2,500	4,532.0
1,398.9	2,550	4,622.0
1,426.7	2,600	4,712.0
1,454.4	2,650	4,802.0
1,482.2	2,700	4,892.0
1,510.0	2,750	4,982.0
1,537.8	2,800	5,072.0
1,565.1	2,850	5,162.0
1,593.3	2,900	5,252.0
1,621.1	2,950	5,342.0
1,648.9	3,000	5,432.0

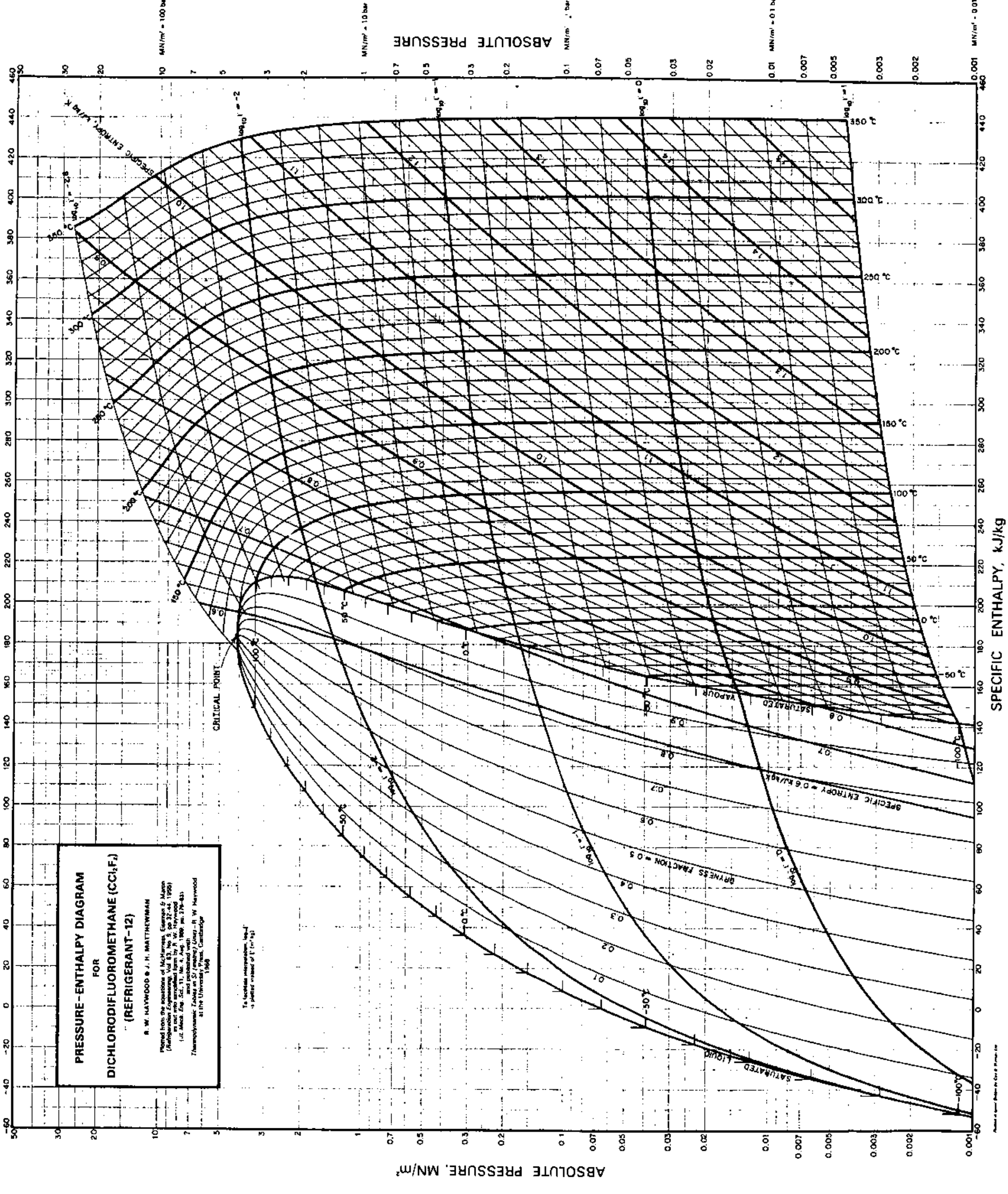
APPENDIX C

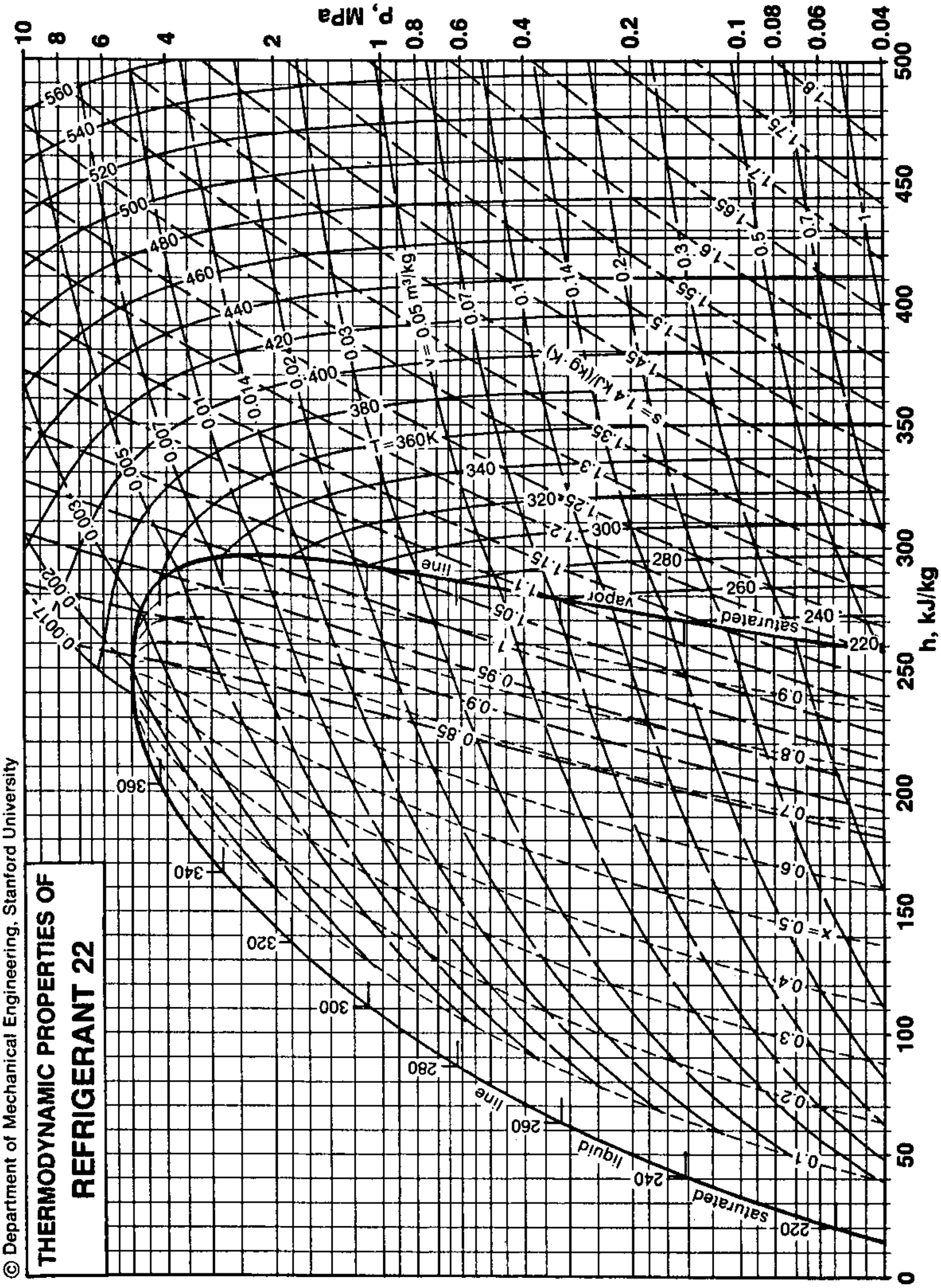
THERMODYNAMIC DATA

Both SI Metric and English unit data are shown. Most is from previous editions. Samples of data from other sources are included.

The H-S diagram for steam, the P-H diagram for Freon 12 and two tables are from the reference shown in the title block. The P-H diagram for Freon 22 is a sample of data from "Thermodynamic Properties in SI" by Reynolds, available from Department of Mechanical Engineering, Stanford University, Stanford, CA 94305.


ENTHALPY—ENTROPY DIAGRAM FOR STEAM
SI
prepared by
D. C. HICKSON AND F. R. TAYLOR
and drawn by
E. A. AYRES
SPECIFIC ENTHALPY kJ/kg
SPECIFIC ENTROPY kJ/kg°K
CONSTANT TEMPERATURE °C
SATURATION LINE
CONSTANT DRYNESS
CONSTANT PRESSURE bar
CRITICAL POINT


PRESSURE-ENTHALPY DIAGRAM
FOR
DICHLORODIFLUOROMETHANE (CCl₂F₂)
(REFRIGERANT-12)
R. W. HAYWOOD • J. H. MATTHEWMAN
Plotted from the equations of McHarness, Eiseman & Martin (Refrigeration Engineering, Vol 63, No 9, pp 32-44, 1955) as put into simplified form by R. W. Haywood (J. Mech. Eng. Sci., 11, No. 4, Aug. 1969, pp. 376-83) and published with Thermodynamic Tables in SI (metric) Units—R. W. Haywood at the University Press, Cambridge 1968
To facilitate interpolation, log₁₀ℓ is plotted instead of ℓ (m³/kg)
ABSOLUTE PRESSURE, MN/m²
ABSOLUTE PRESSURE
SPECIFIC ENTHALPY, kJ/kg
MN/m² = 100 bar
MN/m² = 10 bar
MN/m² = 1 bar
MN/m² = 0.1 bar
MN/m² = 0.01 bar
CRITICAL POINT
SATURATED VAPOUR
SATURATED LIQUID
DRYNESS FRACTION = 0.5
SPECIFIC ENTROPY = 0.6 kJ/kg K
SPECIFIC ENTROPY, kJ/kg K
log₁₀ℓ = -2
log₁₀ℓ = -1
log₁₀ℓ = 0
log₁₀ℓ = 1
350 °C
300 °C
250 °C
200 °C
150 °C
100 °C
50 °C
0 °C
-50 °C
-100 °C
50
30
20
10
7
5
3
2
1
0.7
0.5
0.3
0.2
0.1
0.07
0.05
0.03
0.02
0.01
0.007
0.005
0.003
0.002
0.001
-60
-40
-20
0
20
40
60
80
100
120
140
160
180
200
220
240
260
280
300
320
340
360
380
400
420
440
460



THERMODYNAMIC PROPERTIES OF REFRIGERANT 22

© Department of Mechanical Engineering, Stanford University

Ammonia - NH_3 (Refrigerant 717)

Saturation Values							Superheat $(t-t_s)$			
							50 K		100 K	
t	p_s	v_g	h_f	h_g	s_f	s_g	h	s	h	s
−50	0·4089	2·625	−44·4	1373·3	−0·194	6·159	1479·8	6·592	1585·9	6·948
−45	0·5454	2·005	−22·3	1381·6	−0·096	6·057	1489·3	6·486	1596·1	6·839
−40	0·7177	1·552	0	1390·0	0	5·962	1498·6	6·387	1606·3	6·736
−35	0·9322	1·216	22·3	1397·9	0·095	5·872	1507·9	6·293	1616·3	6·639
−30	1·196	0·9633	44·7	1405·6	0·188	5·785	1517·0	6·203	1626·3	6·547
−28	1·317	0·8809	53·6	1408·5	0·224	5·751	1520·7	6·169	1630·3	6·512
−26	1·447	0·8058	62·6	1411·4	0·261	5·718	1524·3	6·135	1634·2	6·477
−24	1·588	0·7389	71·7	1414·3	0·297	5·686	1527·9	6·103	1638·2	6·444
−22	1·740	0·6783	80·8	1417·3	0·333	5·655	1531·4	6·071	1642·2	6·411
−20	1·902	0·6237	89·8	1420·0	0·368	5·623	1534·8	6·039	1646·0	6·379
−18	2·077	0·5743	98·8	1422·7	0·404	5·593	1538·2	6·008	1650·0	6·347
−16	2·265	0·5296	107·9	1425·3	0·440	5·563	1541·7	5·978	1653·8	6·316
−14	2·465	0·4890	117·0	1427·9	0·475	5·533	1545·1	5·948	1657·7	6·286
−12	2·680	0·4521	126·2	1430·5	0·510	5·504	1548·5	5·919	1661·5	6·256
−10	2·908	0·4185	135·4	1433·0	0·544	5·475	1551·7	5·891	1665·3	6·227
− 8	3·153	0·3879	144·5	1435·3	0·579	5·447	1554·9	5·863	1669·0	6·199
− 6	3·413	0·3599	153·6	1437·6	0·613	5·419	1558·2	5·836	1672·8	6·171
− 4	3·691	0·3344	162·8	1439·9	0·647	5·392	1561·4	5·808	1676·4	6·143
− 2	3·983	0·3110	172·0	1442·2	0·681	5·365	1564·6	5·782	1680·1	6·116
0	4·295	0·2895	181·2	1444·4	0·715	5·340	1567·8	5·756	1683·9	6·090
2	4·625	0·2699	190·4	1446·5	0·749	5·314	1570·9	5·731	1687·5	6·065
4	4·975	0·2517	199·7	1448·5	0·782	5·288	1574·0	5·706	1691·2	6·040
6	5·346	0·2351	209·1	1450·6	0·816	5·263	1577·0	5·682	1694·9	6·015
8	5·736	0·2198	218·5	1452·5	0·849	5·238	1580·1	5·658	1698·4	5·991
10	6·149	0·2056	227·8	1454·3	0·881	5·213	1583·1	5·634	1702·2	5·967
12	6·585	0·1926	237·2	1456·1	0·914	5·189	1586·0	5·611	1705·7	5·943
14	7·045	0·1805	246·6	1457·8	0·947	5·165	1588·9	5·588	1709·1	5·920
16	7·529	0·1693	256·0	1459·5	0·979	5·141	1591·7	5·565	1712·5	5·898
18	8·035	0·1590	265·5	1461·1	1·012	5·118	1594·4	5·543	1715·9	5·876
20	8·570	0·1494	275·1	1462·6	1·044	5·095	1597·2	5·521	1719·3	5·854
22	9·134	0·1405	284·6	1463·9	1·076	5·072	1600·0	5·499	1722·8	5·832
24	9·722	0·1322	294·1	1465·2	1·108	5·049	1602·7	5·478	1726·3	5·811
26	10·34	0·1245	303·7	1466·5	1·140	5·027	1605·3	5·458	1729·6	5·790
28	10·99	0·1173	313·4	1467·8	1·172	5·005	1608·0	5·437	1732·7	5·770
30	11·67	0·1106	323·1	1468·9	1·204	4·984	1610·5	5·417	1735·9	5·750
32	12·37	0·1044	332·8	1469·9	1·235	4·962	1613·0	5·397	1739·3	5·731
34	13·11	0·0986	342·5	1470·8	1·267	4·940	1615·4	5·378	1742·6	5·711
36	13·89	0·0931	352·3	1471·8	1·298	4·919	1617·8	5·358	1745·7	5·692
38	14·70	0·0880	362·1	1472·6	1·329	4·898	1620·1	5·340	1748·7	5·674
40	15·54	0·0833	371·9	1473·3	1·360	4·877	1622·4	5·321	1751·9	5·655
42	16·42	0·0788	381·8	1473·8	1·391	4·856	1624·6	5·302	1755·0	5·637
44	17·34	0·0746	391·8	1474·2	1·422	4·835	1626·8	5·284	1758·0	5·619
46	18·30	0·0706	401·8	1474·5	1·453	4·814	1629·0	5·266	1761·0	5·602
48	19·29	0·0670	411·9	1474·7	1·484	4·793	1631·1	5·248	1764·0	5·584
50	20·33	0·0635	421·9	1474·7	1·515	4·773	1633·1	5·230	1766·8	5·567

Relative molecular mass =17·031; further properties of the liquid are given on p.15.

Dichlorodifluoromethane – CF_2Cl_2 (Refrigerant 12)

Saturation Values							Superheat $(t-t_s)$			
							15 K		30 K	
t	p_s	v_g	h_f	h_g	s_f	s_g	h	s	h	s
−100	0·0118	10·100	−51·84	142·00	−0·2567	0·8628	148·89	0·9019	156·10	0·9428
− 95	0·0181	6·585	−47·56	144·22	−0·2323	0·8442	151·23	0·8830	158·55	0·9195
− 90	0·0284	4·416	−43·28	146·46	−0·2086	0·8274	153·59	0·8649	161·02	0·9010
− 85	0·0424	3·037	−39·00	148·73	−0·1856	0·8122	155·98	0·8493	163·52	0·8851
− 80	0·0617	2·138	−34·72	151·02	−0·1631	0·7985	158·39	0·8351	166·04	0·8706
− 75	0·0879	1·538	−30·43	153·32	−0·1412	0·7861	160·82	0·8226	168·57	0·8578
− 70	0·1227	1·127	−26·13	155·63	−0·1198	0·7749	163·26	0·8110	171·12	0·8459
− 65	0·1680	0·8412	−21·81	157·96	−0·0988	0·7649	165·70	0·8008	173·68	0·8355
− 60	0·2262	0·6379	−17·49	160·29	−0·0783	0·7558	168·15	0·7915	176·26	0·8259
− 55	0·2998	0·4910	−13·14	162·62	−0·0582	0·7475	170·60	0·7830	178·84	0·8172
− 50	0·3915	0·3831	− 8·78	164·95	−0·0384	0·7401	173·07	0·7753	181·43	0·8093
− 45	0·5044	0·3027	− 4·40	167·28	−0·0190	0·7335	175·54	0·7685	184·01	0·8023
− 40	0·6417	0·2419	0	169·60	0	0·7274	178·00	0·7623	186·60	0·7959
− 35	0·8071	0·1954	4·42	171·90	0·0187	0·7219	180·45	0·7568	189·18	0·7902
− 30	1·004	0·1594	8·86	174·20	0·0371	0·7170	182·90	0·7517	191·76	0·7851
− 25	1·237	0·1312	13·33	176·48	0·0552	0·7127	185·33	0·7473	194·33	0·7805
− 20	1·509	0·1088	17·82	178·73	0·0731	0·7087	187·75	0·7432	196·89	0·7764
− 15	1·826	0·0910	22·33	180·97	0·0906	0·7051	190·15	0·7397	199·44	0·7728
− 10	2·191	0·0766	26·87	183·19	0·1080	0·7020	192·53	0·7365	201·97	0·7695
− 5	2·610	0·0650	31·45	185·38	0·1251	0·6991	194·90	0·7336	204·49	0·7666
0	3·086	0·0554	36·05	187·53	0·1420	0·6966	197·25	0·7311	206·99	0·7641
5	3·626	0·0475	40·69	189·66	0·1587	0·6943	199·56	0·7289	209·47	0·7618
10	4·233	0·0409	45·37	191·74	0·1752	0·6921	201·85	0·7268	211·92	0·7598
15	4·914	0·0354	50·10	193·78	0·1915	0·6901	204·10	0·7251	214·35	0·7580
20	5·673	0·0308	54·87	195·78	0·2078	0·6885	206·32	0·7235	216·75	0·7565
25	6·516	0·0269	59·70	197·73	0·2239	0·6869	208·50	0·7220	219·11	0·7552
30	7·449	0·0235	64·59	199·62	0·2399	0·6853	210·63	0·7208	221·44	0·7540
35	8·477	0·0206	69·55	201·45	0·2559	0·6839	212·72	0·7196	223·73	0·7529
40	9·607	0·0182	74·59	203·20	0·2718	0·6825	214·76	0·7185	225·98	0·7519
45	10·84	0·0160	79·71	204·87	0·2877	0·6811	216·74	0·7175	228·18	0·7511
50	12·19	0·0142	84·94	206·45	0·3037	0·6797	218·64	0·7166	230·33	0·7503
55	13·66	0·0125	90·27	207·92	0·3197	0·6782	220·48	0·7156	232·42	0·7496
60	15·26	0·0111	95·74	209·26	0·3358	0·6765	222·23	0·7146	234·45	0·7490
65	16·99	0·00985	101·36	210·46	0·3521	0·6747	223·89	0·7136	236·42	0·7484
70	18·86	0·00873	107·15	211·48	0·3686	0·6726	225·45	0·7125	238·32	0·7477
75	20·88	0·00772	113·15	212·29	0·3854	0·6702	226·89	0·7113	240·13	0·7470
80	23·05	0·00682	119·39	212·83	0·4027	0·6673	228·21	0·7099	241·86	0·7463
85	25·38	0·00601	125·93	213·04	0·4204	0·6636	229·39	0·7084	243·50	0·7455
90	27·89	0·00526	132·84	212·80	0·4389	0·6591	230·43	0·7067	245·03	0·7445
95	30·57	0·00456	140·23	211·94	0·4583	0·6531	231·30	0·7047	246·47	0·7435
100	33·44	0·00390	148·32	210·12	0·4793	0·6449	231·93	0·7023	247·80	0·7424
105	36·51	0·00324	157·52	206·57	0·5028	0·6325	232·22	0·6994	248·97	0·7412
110	39·79	0·00246	169·55	197·99	0·5334	0·6076	232·47	0·6964	250·10	0·7399
112	41·15	0·00179	183·43	183·43	0·5690	0·5690	232·80	0·6958	250·58	0·7394

Specific Heat c_p of Some Gases and Vapours

T	CO_2	CO	H_2	N_2	O_2	H_2O	CH_4	C_2H_4	C_2H_6
175	0·709	1·039	13·12	1·039	0·910	1·850	2·083	1·241	
200	0·735	1·039	13·53	1·039	0·910	1·851	2·087	1·260	
225	0·763	1·039	13·83	1·039	0·911	1·852	2·121	1·316	
250	0·791	1·039	14·05	1·039	0·913	1·855	2·156	1·380	1·535
275	0·819	1·040	14·20	1·039	0·915	1·859	2·191	1·453	1·651
300	0·846	1·040	14·31	1·040	0·918	1·864	2·226	1·535	1·766
325	0·871	1·041	14·38	1·040	0·923	1·871	2·293	1·621	1·878
350	0·895	1·043	14·43	1·041	0·928	1·880	2·365	1·709	1·987
375	0·918	1·045	14·46	1·042	0·934	1·890	2·442	1·799	2·095
400	0·939	1·048	14·48	1·044	0·941	1·901	2·525	1·891	2·199
450	0·978	1·054	14·50	1·049	0·956	1·926	2·703	2·063	2·402
500	1·014	1·064	14·51	1·056	0·972	1·954	2·889	2·227	2·596
550	1·046	1·075	14·53	1·065	0·988	1·984	3·074	2·378	2·782
600	1·075	1·087	14·55	1·075	1·003	2·015	3·256	2·519	2·958
650	1·102	1·100	14·57	1·086	1·017	2·047	3·432	2·649	3·126
700	1·126	1·113	14·60	1·098	1·031	2·080	3·602	2·770	3·286
750	1·148	1·126	14·65	1·110	1·043	2·113	3·766	2·883	3·438
800	1·168	1·139	14·71	1·122	1·054	2·147	3·923	2·989	3·581
850	1·187	1·151	14·77	1·134	1·065	2·182	4·072	3·088	3·717
900	1·204	1·163	14·83	1·146	1·074	2·217	4·214	3·180	3·846
950	1·220	1·174	14·90	1·157	1·082	2·252	4·348	3·266	
1000	1·234	1·185	14·98	1·167	1·090	2·288	4·475	3·347	
1050	1·247	1·194	15·06	1·177	1·097	2·323	4·595	3·423	
1100	1·259	1·203	15·15	1·187	1·103	2·358	4·708	3·494	
1150	1·270	1·212	15·25	1·196	1·109	2·392	4·814	3·561	
1200	1·280	1·220	15·34	1·204	1·115	2·425			
1250	1·290	1·227	15·44	1·212	1·120	2·458			
1300	1·298	1·234	15·54	1·219	1·125	2·490			
1350	1·306	1·240	15·65	1·226	1·130	2·521			
1400	1·313	1·246	15·77	1·232	1·134	2·552			
1500	1·326	1·257	16·02	1·244	1·143	2·609			
1600	1·338	1·267	16·23	1·254	1·151	2·662			
1700	1·348	1·275	16·44	1·263	1·158	2·711			
1800	1·356	1·282	16·64	1·271	1·166	2·756			
1900	1·364	1·288	16·83	1·278	1·173	2·798			
2000	1·371	1·294	17·01	1·284	1·181	2·836			
2100	1·377	1·299	17·18	1·290	1·188	2·872			
2200	1·383	1·304	17·35	1·295	1·195	2·904			
2300	1·388	1·308	17·50	1·300	1·202	2·934			
2400	1·393	1·311	17·65	1·304	1·209	2·962			
2500	1·397	1·315	17·80	1·307	1·216	2·987			
2600	1·401	1·318	17·93	1·311	1·223	3·011			
2700	1·404	1·321	18·06	1·314	1·230	3·033			
2800	1·408	1·324	18·17	1·317	1·236	3·053			
2900	1·411	1·326	18·28	1·320	1·243	3·072			
3000	1·414	1·329	18·39	1·323	1·249	3·090			
3500	1·427	1·339	18·91	1·333	1·276	3·163			
4000	1·437	1·346	19·39	1·342	1·299	3·217			
4500	1·446	1·353	19·83	1·349	1·316	3·258			
5000	1·455	1·359	20·23	1·355	1·328	3·292			
5500	1·465	1·365	20·61	1·362	1·337	3·322			
6000	1·476	1·370	20·96	1·369	1·344	3·350			

T	C_6H_6	C_8H_{18}
250	0·850	1·308
275	0·957	1·484
300	1·060	1·656
325	1·160	1·825
350	1·255	1·979
375	1·347	2·109
400	1·435	2·218
450	1·600	2·403
500	1·752	2·608
550	1·891	2·774
600	2·018	2·924
650	2·134	3·121
700	2·239	3·232
750	2·335	3·349
800	2·422	3·465
850	2·500	3·582
900	2·571	3·673

The specific heats of atomic H, N and O are given with adequate accuracy by $c_p = 2{\cdot}5\ R_0/M$ where M is the relative *atomic* mass.

T = °C. P = kPa

THE PROPERTIES OF SATURATED STEAM

T	P	v_f	v_g	h_f	h_g	s_f	s_g	T	P	v_f	v_g	h_f	h_g	s_f	s_g
40.00	7.38	1.0092	19550.	167.37	2574.3	.5731	8.2586	178.00	957.29	1.1239	202.49	753.91	2774.2	2.1191	6.5971
50.00	12.33	1.0166	12042.	209.19	2591.4	.7054	8.0782	180.00	1001.9	1.1315	194.23	762.62	2774.9	2.1373	6.5832
60.00	19.90	1.0161	7678.7	250.89	2609.4	.8306	7.9109	184.00	1098.1	1.1326	177.60	780.42	2779.2	2.1771	6.5495
65.00	25.03	1.0234	6205.6	271.95	2617.7	.8880	7.8282	186.00	1148.6	1.1367	170.24	789.31	2780.3	2.1961	6.5346
70.00	31.05	1.0191	5045.6	292.57	2626.0	.9625	7.7631	189.00	1227.5	1.1399	159.67	802.57	2783.2	2.2250	6.5107
75.00	38.58	1.0271	4133.2	313.81	2635.1	1.0111	7.6791	191.00	1282.7	1.1427	153.09	811.56	2784.3	2.2438	6.4958
80.00	47.32	1.0290	3409.3	334.66	2643.1	1.0753	7.6131	194.00	1368.7	1.1473	143.81	824.94	2786.8	2.2725	6.4724
83.00	53.40	1.0311	3045.1	347.29	2648.2	1.1097	7.5705	205.00	1723.9	1.1643	115.04	874.80	2793.3	2.3763	6.3901
86.00	60.05	1.0344	2728.0	359.83	2653.0	1.1460	7.5321	215.00	2105.4	1.1824	94.646	887.62	2797.5	2.4698	6.3170
90.00	70.04	1.0378	2362.2	376.63	2659.4	1.1927	7.4798	220.00	2319.2	1.1926	86.098	586.19	2798.9	2.5157	6.2810
100.00	101.28	1.0438	1673.0	418.77	2675.3	1.3063	7.3545	230.00	2796.7	1.2063	71.479	909.56	2801.2	2.6089	6.2095
105.00	120.75	1.0482	1419.4	439.83	2683.1	1.3618	7.2958	235.00	3062.	1.2159	65.284	918.34	2801.4	2.6548	6.1745
110.00	143.21	1.0510	1210.0	460.98	2690.8	1.4178	7.2382	240.00	3345.6	1.2262	59.609	1262.1	2801.1	2.6995	6.1361
115.00	169.01	1.0553	1036.3	482.16	2698.2	1.4727	7.1824	245.00	3651.8	1.2373	54.647	1061.2	2800.6	2.7481	6.1076
120.00	198.42	1.0616	891.50	503.47	2705.5	1.5270	7.1281	250.00	3975.3	1.2518	50.094	1085.0	2798.4	2.7908	6.0655
125.00	232.07	1.0640	770.53	524.54	2712.0	1.5801	7.0764	255.00	4322.1	1.2499	45.859	1110.1	2797.7	2.8395	6.0339
130.00	269.98	1.0707	668.10	545.97	2719.5	1.6336	7.0250	260.00	4692.3	1.2903	42.150	1134.3	2794.0	2.8820	5.9982
135.00	312.94	1.0749	581.89	567.29	2726.0	1.6859	6.9758	265.00	5084.0	1.2818	38.695	1159.4	2791.8	2.9305	5.9619
137.00	331.39	1.0785	551.00	575.91	2729.5	1.7064	6.9555	270.00	5502.5	1.3048	35.579	1184.7	2787.5	2.9747	5.9265
140.00	361.40	1.0788	508.83	588.55	2731.1	1.7388	6.9279	275.00	5946.1	1.3141	32.723	1210.4	2782.9	3.0204	5.8902
145.00	415.34	1.0852	446.07	610.20	2738.7	1.7895	6.8807	277.00	6131.1	1.3205	31.652	1220.7	2780.7	3.0384	5.8757
147.00	438.68	1.0877	423.85	618.73	2740.9	1.8102	6.8620	280.00	6415.0	1.3334	30.113	1236.2	2777.7	3.0674	5.8534
150.00	475.94	1.0907	392.32	631.80	2745.1	1.8409	6.8355	285.00	6913.9	1.3499	27.718	1262.5	2771.4	3.1128	5.8168
153.00	515.36	1.0942	364.10	644.70	2748.4	1.8712	6.8083	290.00	7440.2	1.3719	25.511	1289.2	2764.4	3.1596	5.7795
155.00	543.24	1.0969	346.39	653.38	2751.2	1.8915	6.7907	300.00	8586.3	1.4036	21.618	1344.3	2747.1	3.2536	5.7012
157.00	572.06	1.0976	329.92	662.05	2752.2	1.9115	6.7729	305.00	9208.0	1.4262	19.898	1372.7	2736.6	3.3012	5.6609
160.00	617.90	1.1023	306.75	675.07	2756.4	1.9415	6.7471	310.00	9863.5	1.4647	18.278	1404.7	2721.4	3.3543	5.6139
163.00	666.51	1.1058	285.56	688.12	2759.4	1.9713	6.7213	315.00	10554.	1.4574	16.843	1428.0	2716.1	3.3936	5.5828
165.00	700.56	1.1085	272.40	696.83	2761.5	1.9913	6.7039	335.00	13704.	1.5943	11.833	1557.7	2643.3	3.6017	5.3869
168.00	754.28	1.1124	253.98	709.93	2764.6	2.0210	6.6787	350.00	16526.	1.8019	8.7090	1681.4	2548.0	3.7939	5.1863
170.00	792.94	1.1107	242.25	718.97	2767.7	2.0424	6.6620	370.00	21040.	2.3377	4.8901	1911.4	2320.7	4.1429	4.7793
175.00	892.13	1.1212	216.57	740.62	2771.2	2.0895	6.6217	375.00	22331.	3.3872	2.6537	2154.7	2033.3	4.5143	4.3262

THE PROPERTIES OF SUPERHEATED STEAM

T	P=6.89kPa(Sat.Temp.=38.74°C)			P=137.8kPa(Sat.Temp.=108.87°C)			P=34.5kPa(Sat.Temp.=72.36°C)			P=68.9kPa(Sat.Temp.=89.56°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
95.0	24621.	2678.5	8.5966	1399.8	2773.4	7.4611	4902.1	2674.9	7.8456	2437.1	2669.9	7.5148
150.0	28311.	2783.0	8.8613	1589.4	2882.9	7.6043	5649.1	2781.2	8.1149	2816.8	2778.7	7.7905
205.0	32000.	2888.7	9.0960	1776.7	2992.1	7.9206	6391.4	2887.6	8.3516	3190.1	2886.2	8.0292
260.0	35684.	2996.1	9.3085	1962.7	3102.1	8.1171	7130.5	2995.4	8.5645	3561.5	2994.2	8.2434
315.0	39367.	3105.2	9.5030	2148.1	3213.8	8.2985	7868.4	3104.5	8.7598	3931.1	3103.8	8.4387
370.0	43050.	3215.9	9.6835	2333.0	3327.0	8.4670	8606.4	3215.4	8.9399	4300.7	3215.0	8.6196
425.0	46735.	3328.6	9.8516	2517.7	3442.0	8.6262	9343.7	3328.4	9.1085	4669.7	3327.9	8.7882
480.0	50411.	3443.5	10.009	2702.6	3559.2	8.7757	10080.	3443.2	9.2667	5038.6	3442.8	8.9464
535.0	54095.	3560.3	10.160	2886.7	3678.0	8.9185	10817.	3560.0	9.4167	5407.6	3559.6	9.0964
590.0	57778.	3679.3	10.301	3088.3	3810.4	9.0663	11554.	3679.0	9.5589	5776.0	3678.8	9.2386
650.0	61797.	3811.1	10.450	3457.0	4059.1	9.3211	12357.	3811.1	9.7068	6177.9	3810.8	9.3866
760.0	69158.	4059.8	10.704	3825.5	4316.4	9.5579	13834.	4059.8	9.9612	6915.1	4059.5	9.6414
870.0	76524.	4317.2	10.941				15305.	4316.9	10.198	7651.9	4316.7	9.8777

T	P=101.3kPa(Sat.Temp.=100°C)			P=275.7kPa(Sat.Temp.=130.69°C)			P=413.5kPa(Sat.Temp.-144.84°C)			P=551.3kPa(Sat.Temp.=155.57°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
150.00	1911.1	2776.2	7.6085	691.23	2762.4	7.1207	454.59	2750.4	6.9108	388.81	2863.3	7.0325
205.00	2167.6	2884.6	7.8496	789.31	2876.7	7.3745	522.37	2870.0	7.1765	438.24	2979.3	7.2624
260.00	2421.0	2993.3	8.0642	884.48	2987.9	7.5949	587.01	2983.7	7.4014	486.27	3093.0	7.4656
315.00	2673.1	3103.1	8.2603	978.43	3099.1	7.7935	650.30	3096.1	7.6021	533.55	3206.8	7.6503
370.00	2925.3	3214.5	8.4413	1071.7	3211.4	7.9756	712.97	3209.1	7.7860	580.53	3321.4	7.8218
425.00	3176.2	3327.5	8.6098	1164.7	3325.1	8.1455	775.29	3323.3	7.9562	627.21	3437.4	7.9817
480.00	3427.8	3442.5	8.7685	1257.3	3440.4	8.3045	837.27	3439.0	8.1162	677.97	3566.0	8.1459
535.00	3676.7	3559.4	8.9185	1350.0	3557.7	8.4550	899.21	3556.6	8.2666	720.23	3675.0	8.2760
590.00	3929.0	3678.3	9.0610	1442.7	3677.1	8.5975	960.96	3676.0	8.4095	770.80	3807.6	8.4244
650.00	4203.3	3810.6	9.2090	1543.2	3809.5	8.7460	1028.2	3808.5	8.5580	63.38	4057.0	8.6755
760.00	4705.2	4059.3	9.4634	1728.0	4058.4	9.0008	1551.5	4057.7	8.8128	955.78	4314.8	8.9166
870.00	5206.4	4316.6	9.7002	1912.2	4316.0	9.2375	1274.8	4315.3	9.0500			

T	P=689.2kPa(Sat.Temp.=164.34°C)			P=827.0kPa(Sat.Temp.=171.81°C			P=964.8kPa(Sat.Temp.=178.34°C)			P=1102.7kPa(Sat.Temp.=184.18°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
205.00	308.62	2856.1	6.9183	255.13	2848.7	6.8217	217.82	2841.1	6.7381	188.07	2833.0	6.6632
260.00	348.90	2974.7	7.1531	289.42	2970.2	7.0627	246.84	2965.6	6.9848	214.94	2960.7	6.9162
315.00	387.79	3089.8	7.3589	322.12	3086.5	7.2705	275.17	3083.5	7.1951	240.03	3080.2	7.1294
370.00	425.92	3204.4	7.5448	354.13	3201.9	7.4581	303.64	3199.5	7.3839	264.46	3197.0	7.3195
425.00	463.70	3319.5	7.7167	386.79	3317.6	7.6304	329.92	3315.8	7.5576	288.38	3313.7	7.4939
480.00	501.16	3436.0	7.8770	417.13	3434.3	7.7916	358.48	3432.9	7.7187	312.12	3431.3	7.6560
540.00	541.88	3564.8	8.0416	451.18	3563.4	7.9567	386.37	3562.3	7.8842	337.81	3560.9	7.8215
590.00	575.76	3673.9	8.1717	479.44	3672.7	8.0867	410.65	3671.5	8.0147	359.11	3670.3	7.9523
650.00	616.29	3806.4	8.3206	513.28	3805.5	8.2356	440.71	3804.6	8.1640	384.53	3803.7	8.1016
760.00	690.45	4056.3	8.5762	575.21	4055.6	8.4912	492.87	4054.7	8.4197	431.13	4054.0	8.3577
870.00	764.50	4314.1	8.8134	636.95	4313.7	8.7288	547.86	4312.9	8.6572	477.49	4312.5	8.5953

T	P=1240.5kPa(Sat.Temp.=189.48°C)			P=1378.3kPa(Sat.Temp.=194.33°C)			P=1516.2kPa(Sat.Temp.=198.81°C)			P=1654.0kPa(Sat.Temp.=202.98°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
205.00	166.63	2824.7	6.5951	147.64	2816.1	6.5628	132.89	2807.3	6.4731	120.55	2798.1	6.4171
260.00	190.03	2955.8	6.8550	170.18	2950.9	6.7994	153.88	2945.8	6.7479	140.28	2940.7	6.7001
315.00	212.72	3076.7	7.0711	190.83	3073.4	6.8901	172.86	3070.2	6.9698	157.96	3066.7	6.9254
370.00	234.54	3194.6	7.2625	210.61	3192.0	7.4687	191.04	3189.7	7.1641	174.71	3187.1	7.1209
425.00	255.94	3311.8	7.4373	229.97	3309.9	4.7009	208.72	3308.1	7.3406	191.04	3306.2	7.2983
480.00	277.07	3429.6	7.5999	249.06	3428.2	7.5500	226.17	3426.6	7.5043	207.05	3425.0	7.4624
540.00	300.00	3559.7	7.7658	239.76	3558.3	7.7160	244.98	3557.2	7.6712	224.38	3555.8	7.6298
590.00	318.93	3669.4	7.8970	286.84	3668.2	7.8472	260.60	3667.0	7.8024	238.71	3666.1	7.7613
650.00	341.65	3802.7	8.0464	307.34	3801.8	7.9970	279.21	3800.9	7.9522	255.83	3799.9	7.9112
760.00	383.06	4053.3	8.3028	344.66	4052.6	8.2534	313.20	4051.9	8.2091	286.98	4051.2	8.1684
870.00	424.33	4311.8	8.5405	381.86	4311.3	8.4914	347.08	4310.6	8.4471	318.06	4310.2	8.4065

h = kJ/kg, v = cm^3/g and S = kJ/kg·K

Datum - Sat. liquid h = 0 at 0°C.

T = °C,P = kPa

THE PROPERTIES OF SATURATED METHANE

T	P	v_f	v_g	h_f	h_g	s_f	s_g
-170.00	47.43	2.2926	1105.8	11.38	536.39	0.1068	5.1964
-165.00	74.45	2.3290	726.41	28.67	544.99	0.2626	5.0370
-160.00	114.74	2.3694	488.69	46.75	533.60	0.4093	4.8890
-155.00	169.41	2.4105	341.06	63.60	561.24	0.5490	4.7611
-150.00	241.21	2.4538	247.13	81.57	568.99	0.6924	4.6506
-145.00	330.38	2.5023	183.92	99.70	575.78	0.8382	4.5536
-140.00	444.52	2.5545	139.21	117.67	581.86	0.9768	4.4631
-135.00	586.36	2.6109	107.15	136.24	587.37	1.1120	4.3776
-130.00	755.82	2.6735	84.28	155.63	591.73	1.2511	4.2977
-125.00	957.07	2.7421	66.714	175.37	595.01	1.3886	4.2245
-120.00	1199.8	2.8131	53.195	195.01	597.09	1.5164	4.1148
-115.00	1481.9	2.9419	42.800	215.98	598.00	1.6552	4.1617
-110.00	1801.0	2.9119	34.739	238.04	597.51	1.7852	3.8299
-105.00	2179.3	3.5151	27.765	266.23	289.82	1.9521	4.0774
-100.00	2608.1	3.1466	22.776	284.01	592.06	2.0478	3.9673
-95.00	3088.7	3.3192	18.295	313.55	582.64	2.2032	3.7067
-90.00	3632.0	3.7450	14.471	344.42	571.86	2.3681	3.6099
-85.00	4248.0	4.1518	10.702	389.03	546.58	2.5962	3.4361

THE PROPERTIES OF SUPERHEATED METHANE

T	P=68.92kPa (Sat.Temp.=-165.89°C)			P=137.84kPa (Sat.Temp =-157.67°C)			P=206.75kPa (Sat.Temp.=-152.22°C)			P=275.67kPa (Sat.Temp.=-147.94°C)			P=413.51kPa (Sat.Temp.=-141.22°C)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
-160.00	827.88	555.99	5.1754												
-150.00	907.30	577.52	5.3579	444.03	574.15	4.9740	290.45	570.49	4.7409						
-140.00	985.11	598.84	5.5324	484.44	596.05	5.1552	317.76	592.79	4.9266	234.10	589.77	4.7583	151.14	583.95	4.5113
-130.00	1060.3	620.26	5.6957	523.96	617.64	5.3216	344.27	614.83	5.0974	254.55	612.14	4.9332	165.57	607.03	4.6949
-120.00	1145.7	640.09	5.8390	566.03	637.78	5.4697	376.38	634.32	5.2477	278.19	633.78	5.0870	179.90	628.56	4.8529
-105.00	1256.0	672.06	6.0412	623.20	669.97	5.6741	414.69	667.12	5.4549	307.56	666.37	5.2977	200.99	662.08	5.0699
-95.00	1332.1	694.28	6.1663	661.09	652.44	5.8003	436.91	690.86	5.5821	325.93	687.46	5.4270	214.87	684.99	5.2024
-85.00	1407.7	715.23	6.2783	700.04	713.37	5.9135	463.30	711.62	5.6969	345.85	708.45	5.5425	228.38	706.80	5.3201
-80.00	1446.3	725.58	6.3311	719.98	723.63	5.9672	477.11	721.65	5.7513	356.29	657.67	5.5971	235.18	717.65	5.3756
-75.00	1488.5	736.37	6.3844	739.13	734.33	6.0210	489.80	732.50	5.8055	366.22	506.28	5.6520	241.94	728.89	5.4311
-60.00	1606.0	767.77	6.5374	795.52	766.40	6.1739	528.57	764.88	5.9601	394.85	585.51	5.8071	261.34	761.20	5.5880
-50.00	1667.7	788.85	6.6339	834.03	787.52	6.2712	553.91	786.22	6.0577	414.00	601.63	5.9053	274.18	782.86	5.6873
-40.00	1749.2	810.00	6.7265	872.12	808.84	6.3639	579.33	807.67	6.1508	433.25	806.74	5.9988	287.17	804.42	5.7824
-30.00	1845.1	831.22	6.8161	910.04	830.14	6.4532	604.97	828.94	6.2405	452.98	828.04	6.0890	300.46	826.05	5.8734
-20.00	1902.0	852.95	6.9022	946.98	851.79	6.5408	631.62	850.71	6.3283	472.08	849.63	6.1769	312.21	848.17	5.9614
-5.00	2014.7	884.79	7.0272	1004.8	884.08	6.6656	669.07	883.08	6.4530	500.72	882.08	6.3022	332.83	880.40	6.0880
5.00	2089.0	906.85	7.1069	1042.2	905.89	6.7456	694.11	904.99	6.5333	519.87	904.07	6.3827	345.31	902.67	6.1687
15.00	2163.7	928.85	7.1841	1080.5	927.89	6.8228	718.77	927.21	6.6109	539.06	926.47	6.4603	358.20	925.08	6.2466
25.00	2241.1	950.73	7.2603	1118.3	950.08	6.8988	745.33	949.23	6.6875	557.28	948.81	6.5362	371.45	947.56	6.3236
40.00	2354.0	985.02	7.3730	1175.7	984.39	7.0115	782.19	983.54	6.8001	585.94	983.09	6.6496	390.77	982.16	6.4369
50.00	2429.4	1007.6	7.4448	1213.4	1006.9	7.0836	808.40	1006.2	6.8726	605.56	1005.7	6.7219	403.38	1004.7	6.5096
60.00	2504.6	1030.7	7.5136	1251.7	1030.0	7.1527	833.41	1029.5	6.9417	624.90	1028.8	6.7914	415.77	1027.9	6.5791
70.00	2579.7	1054.3	7.5816	1290.0	1053.9	7.2206	857.74	1053.2	7.0093	643.40	1052.8	6.8594	428.47	1051.8	6.6483
80.00	2655.5	1077.2	7.6504	1327.6	1076.8	7.2899	884.48	1076.0	7.0782	662.78	1075.5	6.9284	442.72	1074.6	6.7008
95.00	2767.5	1113.9	7.7501	1384.1	1113.5	7.3894	922.15	1113.0	7.1783	691.64	1112.5	7.0386	460.95	1111.6	6.8013
105.00	2842.8	1138.4	7.8157	1421.7	1137.9	7.4553	947.77	1137.4	7.2442	710.17	1137.0	7.0943	472.34	1136.0	6.8920
115.00	2918.0	1163.5	7.8798	1459.4	1163.0	7.5193	973.09	1162.5	7.3083	729.22	1162.1	7.1584	484.20	1161.1	6.9577
125.00	2993.6	1188.7	7.9436	1496.8	1188.4	7.5829	998.16	1187.9	7.3719	748.97	1187.5	7.2227	498.30	1186.7	6.9526
140.00	3106.5	1227.0	8.0385	1553.2	1226.7	7.6779	1035.2	1226.3	7.4669	776.70	1225.8	7.3176	516.83	1225.0	7.1057
150.00	3181.4	1253.1	8.1003	1591.3	1252.6	7.7398	1060.5	1252.2	7.5288	795.49	1251.7	7.3794	529.76	1251.0	7.1674
160.00	3256.2	1279.5	8.1613	1628.7	1279.1	7.8013	1085.6	1278.6	7.5902	814.06	1278.1	7.4404	542.50	1277.4	7.2289
170.00	3331.7	1306.3	8.2222	1665.8	1306.0	7.8623	1110.4	1305.5	7.6512	833.15	1305.0	7.5012	554.79	1304.4	7.2902
180.00	3406.3	1333.6	8.2830	1703.6	1333.4	7.9226	1135.1	1333.3	7.7120	851.98	1332.7	7.5638	567.63	1331.9	7.3512
195.00	3519.6	1375.0	8.3722	1760.3	1374.7	8.0128	1173.4	1374.5	7.8012	880.02	1373.9	7.6517	586.35	1373.5	7.4404
205.00	3594.4	1403.0	8.4322	1797.5	1402.7	8.0718	1198.1	1402.4	7.8611	898.78	1402.0	7.7117	598.74	1401.5	7.5002
215.00	3669.7	1431.2	8.4917	1835.2	1431.0	8.1313	1223.4	1430.5	7.9207	917.90	1430.3	7.7712	611.69	1429.8	7.5597
225.00	3745.0	1460.0	8.5506	1872.8	1459.8	8.1900	1248.8	1459.5	7.9795	936.47	1459.3	7.8300	624.67	1458.9	7.6190
240.00	3857.9	1504.0	8.6382	1929.3	1503.8	8.2782	1285.9	1503.5	8.0671	964.33	1503.3	7.9176	643.24	1502.8	7.7066
250.00	3933.2	1533.5	8.6961	1966.9	1533.3	8.3360	1311.3	1533.1	8.1250	983.50	1532.9	7.9750	655.63	1532.4	7.7646
260.00	4007.9	1563.7	8.7533	2003.9	1563.5	8.3933	1336.0	1563.5	8.1823	1002.0	1563.3	8.0328	667.98	1562.8	7.8218

T	P=551.34kPa (Sat.Temp.=-136.11°C)			P=689.18kPa (Sat.Temp.=-131.94°C)		
	v	h	s	v	h	s
-160.00						
-150.00						
-140.00						
-130.00	120.78	600.79	4.5612	93.878	594.74	4.3656
-120.00	131.26	623.32	4.6783	101.83	618.29	4.5371
-105.00	147.97	657.55	4.9027	116.08	653.63	4.7679
-95.00	159.22	681.31	5.0385	125.68	677.24	4.9072
-85.00	169.60	703.31	5.1588	134.15	699.74	5.0301
-80.00	174.50	714.03	5.2152	138.13	710.84	5.0874
-75.00	179.49	725.15	5.2718	142.39	722.08	5.1447
-60.00	194.44	758.44	5.4315	154.41	755.66	5.3067
-50.00	204.27	780.41	5.5316	162.45	777.93	5.4082
-40.00	214.25	802.33	5.6271	170.37	800.00	5.5048
-30.00	224.27	824.05	5.7188	178.35	821.83	5.5972
-20.00	233.48	845.99	5.8077	186.32	844.50	5.6872
-5.00	248.73	878.94	5.9348	198.25	877.43	5.8146
5.00	258.42	901.30	6.0158	205.98	899.62	5.8961
15.00	268.00	923.72	6.0942	213.63	922.04	5.9744
25.00	277.22	945.81	6.1717	222.35	944.64	6.0520
40.00	292.41	980.41	6.2880	234.07	979.22	6.1658
50.00	301.95	1003.2	6.3576	241.56	1002.1	6.2388
60.00	311.51	1026.5	6.4272	248.46	1025.4	6.3087
70.00	321.32	1050.4	6.4957	256.67	1049.1	6.3770
80.00	329.78	1073.3	6.5642	263.30	1072.5	6.4465
95.00	344.65	1110.3	6.6653	275.69	1109.5	6.5473
105.00	353.85	1134.9	6.7314	282.55	1134.2	6.6138
115.00	363.88	1160.3	6.7962	290.38	1159.5	6.6786
125.00	374.01	1185.4	6.8604	298.58	1185.0	6.7426
140.00	387.27	1224.1	6.9554	309.49	1223.3	6.8381
150.00	397.73	1250.1	7.0171	317.49	1249.4	6.8999
160.00	407.03	1276.7	7.0786	325.25	1275.8	6.9614
170.00	415.97	1303.8	7.1401	332.73	1302.7	7.0228
180.00	425.71	1330.9	7.2005	341.50	1330.4	7.0837
195.00	440.36	1372.6	7.2903	351.43	1371.7	7.1735
205.00	449.29	1400.9	7.3504	359.44	1400.0	7.2336
215.00	458.90	1429.4	7.4099	367.29	1428.4	7.2931
225.00	468.58	1458.2	7.4687	374.60	1457.4	7.3518
240.00	482.19	1502.4	7.5568	385.55	1501.7	7.4400
250.00	492.48	1532.2	7.6146	392.87	1531.2	7.4978
260.00	501.30	1562.3	7.6723	400.79	1561.6	7.5555

T	P=1033.77kPa (Sat.Temp.=-123.33°C)			P=1378.4kPa (Sat.Temp.=-116.78°C)		
	v	h	s	v	h	s
-115.00	67.538	618.16	4.3507			
-105.00	74.382	641.32	4.5090	52.594	629.52	4.3121
-95.00	80.379	667.57	4.6587	57.937	656.72	4.4701
-85.00	86.482	691.18	4.7883	62.802	681.36	4.6054
-80.00	89.462	702.55	4.8485	65.061	693.33	4.6680
-75.00	92.300	714.43	4.9085	67.289	705.84	4.7301
-60.00	100.89	749.00	5.0 55	74.043	741.60	4.9040
-50.00	106.46	771.58	5.1794	78.361	765.14	5.0111
-40.00	111.93	794.19	5.2779	82.654	788.37	5.1125
-30.00	117.34	816.66	5.3723	86.839	811.37	5.2091
-15.00	125.64	850.22	5.5074	93.224	845.57	5.3464
-5.00	131.00	872.57	5.5935	97.308	868.50	5.4338
5.00	136.18	895.51	5.6759	101.37	891.52	5.5173
15.00	141.35	918.34	5.7553	105.36	914.55	5.5975
25.00	146.67	940.93	5.8343	109.30	937.68	5.6769
40.00	154.65	976.12	5.9492	115.45	972.85	5.7930
50.00	160.08	999.29	6.0227	119.42	996.35	5.8670
60.00	165.3	1022.8	6.0926	123.36	1020.0	5.9377
70.00	170.45	1046.9	6.1618	127.38	1044.0	6.0072
80.00	175.46	1070.0	6.2316	130.99	1067.7	6.0776
95.00	183.28	1107.1	6.3331	137.10	1104.9	6.1795
105.00	188.33	1131.9	6.3994	140.87	1129.8	6.2466
115.00	193.57	1157.7	6.4642	144.79	1155.3	6.3118
125.00	199.13	1183.1	6.5289	148.73	1181.1	6.3763
140.00	206.66	1222.1	6.6245	154.53	1220.1	6.4723
150.00	211.35	1248.3	6.6864	158.43	1246.4	6.5350
160.00	215.94	1274.9	6.7479	162.25	1273.0	6.5967
170.00	221.53	1301.9	6.8097	166.07	1300.0	6.6582
180.00	227.29	1329.2	6.8706	169.94	1327.4	6.7190
190.00	232.42	1357.0	6.9304	173.83	1355.1	6.7790

h = kJ/kg, v = cm^3/g and S = kJ/kg·K

Datum - Sat. liquid h and s = 0 at -173.3°C

THE PROPERTIES OF SUPERHEATED METHANE (continued)

T	P=2067.5kPa (Sat.Temp.=-106.39°C)			P=3445.90kPa (Sat.Temp.=-91.61°C)			P=5513.4kPa			P=6891.8kPa			P=10337.7kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
-115.00															
-105.00															
-95.00	34.808	632.20	4.1547												
-85.00	38.765	660.60	4.3155	18.498	606.89	3.8389									
-80.00	40.512	674.07	4.3895	20.303	627.27	3.9425									
-75.00	42.261	687.44	4.4599	21.834	646.32	4.0379	10.103	576.99	3.5390						
-60.00	47.323	726.35	4.6481	25.659	693.70	4.2770	12.843	633.43	3.8147	8.4124	570.54	3.4708			
-50.00	50.413	751.35	4.7627	27.870	722.50	4.4103	14.929	673.60	4.0017	10.513	628.14	3.7376	5.5479	528.01	3.2672
-40.00	53.438	775.81	4.8680	30.028	750.00	4.5297	16.693	707.67	4.1537	12.217	672.56	3.9343	6.6735	589.07	3.5056
-30.00	56.379	799.87	4.9665	32.125	776.44	4.6399	18.275	739.25	4.2882	13.649	711.32	4.0910	7.8017	638.03	3.6888
-15.00	60.726	835.58	5.1106	34.898	815.56	4.7944	20.362	784.61	4.4672	15.530	762.66	4.2905	9.3276	704.23	3.9297
-5.00	63.618	859.71	5.2004	36.781	841.34	4.8915	21.710	813.66	4.5748	16.736	794.03	4.4067	10.249	744.36	4.0652
5.00	66.452	883.35	5.2864	38.617	866.97	4.9835	22.975	842.02	4.6740	17.834	825.10	4.5124	11.094	782.30	4.1859
15.00	69.211	907.02	5.3689	40.4 0	892.28	5.0708	24.209	869.66	4.7675	18.865	854.89	4.6107	11.893	817.65	4.2963
25.00	71.981	931.03	5.4492	42.180	917.61	5.1560	25.479	896.71	4.8580	19.890	882.26	4.7048	12.669	850.42	8.4023
40.00	76.198	966.53	5.5682	44.798	954.63	5.2786	27.241	936.29	4.9882	21.391	924.49	4.8400	13.775	896.23	4.5509
50.00	78.900	990.36	5.6437	46.526	979.21	5.3570	28.409	961.83	5.0699	22.407	950.48	4.9250	14.473	925.21	4.6446
60.00	81.593	1014.4	5.7158	48.257	1003.7	5.4299	29.528	987.44	5.1473	23.348	976.98	5.0049	15.164	953.49	4.7319
70.00	84.306	1038.7	5.7861	49.933	1028.5	5.5003	30.651	1013.3	5.2223	24.211	1003.8	5.0823	15.837	981.50	4.8152
80.00	86.912	1062.3	5.8572	51.640	1053.0	5.5731	31.751	1038.7	5.2964	25.246	1029.2	5.1602	16.483	1008.4	4.8988
95.00	90.921	1100.2	5.9602	54.084	1091.1	5.6768	33.377	1078.3	5.4049	26.502	1069.6	5.2711	17.462	1050.0	5.0154
105.00	93.548	1125.7	6.0277	55.714	1116.9	5.7446	34.457	1104.8	5.4752	27.386	1096.7	5.3434	18.089	1077.6	5.0911
115.00	96.247	1151.6	6.0933	57.344	1142.0	5.8109	35.579	1131.3	5.5438	28.281	1124.1	5.4131	18.707	1105.4	5.1638
125.00	98.865	1177.1	6.1581	58.960	1183.4	5.8771	36.668	1157.9	5.6113	29.220	1151.1	5.4818	19.325	1133.1	5.2348
140.00	102.82	1216.1	6.2553	61.396	1223.4	5.9749	38.181	1198.4	5.7118	30.470	1191.6	5.8535	20.242	1175.2	5.3394
150.00	105.41	1242.5	6.3181	63.050	1231.9	6.0385	39.264	1225.6	5.7770	31.375	1219.1	5.6489	20.828	1203.6	5.4067
160.00	108.00	1269.3	6.3803	64.675	1242.3	6.1014	40.328	1253.0	5.8410	32.212	1246.7	5.7133	21.407	1232.1	5.4730
170.00	110.58	1296.5	6.4421	66.247	1288.4	6.1644	41.340	1280.8	5.9047	32.999	1274.6	5.7778	21.996	1260.8	5.5390
180.00	113.22	1323.9	6.5036	67.865	1322.0	6.2264	42.348	1309.0	5.9678	33.829	1302.9	5.8413	22.592	1289.6	5.6040
190.00	115.85	1351.7	6.5643	69.494	1335.8	6.2874	43.370	1337.5	6.0297	34.699	1331.7	5.9034	23.169	1318.8	5.6680

h = kJ/kg, v = cm^3/g and s = kJ/kg·K

T = °C, P = kPa

THE PROPERTIES OF SATURATED PROPANE

T	P	v_f	v_g	h_f	h_g	s_f	s_g	T	P	v_f	v_g	h_f	h_g	s_f	s_g
-62.22	33.94	1.6543	1011.3	378.14	823.26	3.6819	5.7912	15.56	736.73	1.9665	61.429	557.21	912.09	4.4007	5.6275
-61.11	41.35	1.6606	955.15	380.47	824.65	3.6932	5.7870	21.11	856.65	2.0033	53.313	571.40	917.21	4.4481	5.6216
-60.00	43.56	1.6606	911.45	382.79	826.05	3.7041	5.7828	26.67	989.66	2.0408	46.509	585.81	921.86	4.4954	5.6158
-56.67	51.55	1.6731	780.35	389.77	830.23	3.7376	5.7698	32.22	1137.2	2.0782	40.141	600.47	926.28	4.5427	5.6103
-51.11	67.40	1.6874	609.92	401.63	837.21	3.7932	5.7527	37.78	1300.5	2.1163	34.835	615.35	930.70	4.5900	5.6053
-45.56	86.84	1.7062	482.57	413.49	843.72	3.8468	5.7368	43.33	1480.4	2.1550	30.402	630.47	934.65	4.6390	5.6011
-40.00	110.27	1.7249	384.56	425.58	850.47	3.9000	5.7234	48.89	1677.5	2.2049	26.594	646.51	939.07	4.6871	5.5969
-34.44	139.08	1.7442	313.39	438.14	857.21	3.9528	5.7108	54.44	1891.8	2.2549	23.098	663.26	942.79	4.7353	5.5919
-28.89	172.64	1.7642	253.46	450.70	863.95	4.0059	5.6982	60.00	2125.4	2.3111	19.977	680.70	946.51	4.7855	5.5881
-23.33	213.30	1.7848	207.88	463.72	870.70	4.0570	5.6865	65.56	2380.4	2.3829	17.355	698.14	949.30	4.8366	5.5793
-17.78	260.58	1.8060	171.05	476.74	877.21	4.1081	5.6752	71.11	2653.3	2.4734	14.983	717.21	950.70	4.8902	5.5697
-12.22	315.99	1.8291	143.58	490.00	883.72	4.1583	5.6652	76.67	2935.9	2.5795	12.985	738.37	950.23	4.9471	5.5567
-6.67	379.05	1.8541	120.49	503.72	889.77	4.2077	5.6564	82.22	3261.2	2.7262	11.237	761.63	947.91	5.0116	5.5362
-1.11	452.79	1.8797	99.885	516.98	895.58	4.2567	5.6484	86.67	3432.8	2.8985	9.6763	783.49	943.95	5.0685	5.5136
4.44	536.18	1.9072	83.029	530.00	901.16	4.3053	5.6409	87.78	3607.2	2.9416	9.3017	788.84	940.93	5.0828	5.5082
10.00	630.60	1.9359	71.168	543.72	906.74	4.3534	5.6338	93.33	3962.8	3.2525	7.0543	822.09	926.28	5.1749	5.4596

THE PROPERTIES OF SUPERHEATED PROPANE

T	P = 50.65kPa (Sat.Temp.=-57.06°C)			P = 84.36kPa (Sat.Tem.=-46.11°C)			P = 101.28kPa (Sat.Temp.=-42.06°C)			P = 137.84kPa (Sat.Temp.=-34.61°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
-50.00	822.56	839.39	5.8130									
-40.00	860.88	853.02	5.8711	510.78	851.63	5.7761	422.95	850.70	5.7401			
-30.00	899.16	866.89	5.9293	533.79	865.81	5.8334	442.48	865.16	5.7988	322.20	863.95	5.7372
-15.00	955.54	888.34	6.0152	568.29	887.46	5.9194	471.80	887.06	5.8822	343.56	886.33	5.8241
0.00	1013.4	911.14	6.1011	602.63	910.37	6.0051	500.80	910.01	5.9900	365.08	908.96	5.9104
10.00	1050.6	927.09	6.1585	625.90	925.96	6.9625	520.19	925.82	5.9681	379.56	925.15	5.9669
20.00	1089.2	942.52	6.2154	648.64	941.65	6.1197	539.28	941.72	6.1909	393.79	940.92	6.0236
30.00	1126.8	961.12	6.2730	672.20	960.62	6.1755	558.49	958.89	6.0170	408.14	957.93	6.0800
40.00	1164.4	973.74	6.3294	694.70	974.31	6.2316	578.11	975.81	6.1281	422.44	974.90	6.1354
50.00	1202.9	994.70	6.3847	717.35	993.61	6.2872	596.88	993.13	6.2527	436.63	992.46	6.1904
60.00	1241.1	1012.1	6.4393	740.39	1013.7	6.3422	616.10	1011.2	6.3070	450.79	1010.7	6.2454
70.00	1278.7	1024.1	6.4936	763.43	1030.3	6.3968	635.36	1029.8	6.3634	464.95	1029.3	6.3005
80.00	1316.8	1046.8	6.5478	786.46	1048.6	6.4511	652.54	1048.9	6.3860	479.11	1048.3	6.3555
90.00	1355.0	1074.5	6.6021	809.52	1069.8	6.5053	669.51	1068.7	6.4262	493.27	1067.7	6.4105

T	P = 206.75kPa (Sat.Temp.=-24.18°C)			P = 275.67kPa (Sat.Temp.=-16.17°C)			P = 413.51kPa (Sat.Temp.=4.00°C)			P = 551.34kPa (Sat.Temp.=5.38°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
-15.00	224.94	884.19	5.7428									
0.00	239.75	907.14	5.8281	176.96	905.01	5.7679						
10.00	249.65	923.21	5.8844	184.58	921.12	5.8257	119.32	916.82	5.7395			
20.00	259.52	939.23	5.9413	192.25	937.56	5.8826	124.65	933.37	5.8003	90.822	928.88	5.7316
30.00	269.22	956.20	5.9973	199.64	954.59	5.9388	129.92	950.76	5.8591	95.020	947.11	5.7910
40.00	278.80	973.74	6.0531	206.95	971.73	5.9956	135.03	968.71	5.9171	99.111	964.84	5.8494
50.00	288.35	991.26	6.1089	214.25	989.45	6.0520	140.09	986.63	5.9741	103.06	982.98	5.9073
60.00	297.91	1009.3	6.1646	221.56	1007.9	6.1077	145.14	1005.1	6.0307	107.00	1002.1	5.9654
70.00	307.46	1028.1	6.2204	228.86	1026.9	6.1635	150.20	1024.4	6.0872	110.94	1021.8	6.0234
80.00	317.01	1047.3	6.2762	236.16	1046.3	6.2193	155.26	1044.0	6.1434	114.77	1041.6	6.0808
90.00	326.56	1067.0	6.3319	243.47	1066.1	6.2746	160.25	1064.1	6.1993	118.57	1062.0	6.1375

T	P = 689.16kPa (Sat.Temp.=13.12°C)			P = 895.93kPa (Sat.Temp.=22.89°C)			P = 1102.7kPa (Sat.Temp.=30.95°C)			P = 1309.4kPa (Sat.Temp.=38.06°C)			P = 1516.2kPa (Sat.Temp.=44.36°C)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
20.00	70.164	923.71	5.6734												
30.00	73.857	942.53	5.7417	53.892	934.03	5.6691									
40.00	77.305	960.77	5.7942	57.207	954.07	5.7322	44.155	945.57	5.6760						
50.00	80.897	979.58	5.8530	60.159	973.72	5.7939	46.886	966.59	5.7424	37.705	959.75	5.6934	30.924	951.05	5.6438
60.00	84.090	999.07	5.9159	62.802	993.72	5.8561	49.368	987.91	5.8058	40.047	981.63	5.7606	33.174	974.88	5.7150
70.00	87.050	1018.9	5.9748	65.386	1014.5	5.9176	51.683	1009.0	5.8676	42.173	1003.1	5.8251	35.205	997.00	5.7817
80.00	90.373	1039.2	6.0328	67.974	1035.0	5.9780	53.904	1030.4	5.9295	44.171	1025.3	5.8880	37.049	1019.7	5.8470
90.00	93.717	1059.9	6.0918	70.523	1055.9	6.0377	56.058	1051.9	5.9908	46.048	1047.2	5.9486	38.786	1042.0	5.9090

Datum - h = 0 and s = 0 for solid at absolute zero

T = °C. P = kPa

THE PROPERTIES OF SATURATED AMMONIA

T	P	v_f	v_g	h_f	h_g	s_f	s_g
-50.00	40.87	1.4246	2625.1	-44.42	1373.0	-0.1943	6.1600
-45.00	66.05	1.4362	1919.9	-22.14	1338.4	-0.3063	5.9975
-40.00	71.74	1.4489	1552.0	0.00	1389.8	-0.0000	5.9628
-35.00	92.43	1.5175	1222.2	18.23	1401.7	0.1105	5.8760
-30.00	119.54	1.4749	963.89	45.21	1405.3	0.1874	5.7856
-28.00	131.59	1.4821	880.53	53.50	1408.3	0.2246	5.7521
-26.00	144.65	1.4860	805.50	62.66	1411.2	0.2605	5.7191
-24.00	158.69	1.4923	738.66	71.77	1414.2	0.2970	5.6870
-22.00	173.82	1.4977	678.42	80.75	1417.0	0.3329	5.6554
-20.00	190.14	1.5033	623.72	89.77	1419.8	0.3684	5.6241
-18.00	207.64	1.5102	574.28	98.78	1422.5	0.4043	5.5936
-16.00	226.33	1.5151	529.57	107.75	1425.0	0.4397	5.5637
-14.00	246.50	1.5212	488.95	116.88	1427.7	0.4749	5.5343
-12.00	267.57	1.5274	452.08	126.24	1430.4	0.5095	5.5049
-10.00	291.20	1.5335	418.43	135.17	1432.6	0.5444	5.4766
-8.00	314.68	1.5397	387.86	144.32	1435.1	0.5788	5.4483
-6.00	340.80	1.5460	359.85	153.54	1437.7	0.6132	5.4210
-4.00	368.78	1.5521	334.31	162.74	1439.8	0.6473	5.3937
-2.00	398.11	1.5590	310.95	171.95	1441.9	0.6812	5.3666
0.00	429.29	1.5663	289.48	181.16	1444.2	0.7151	5.3403
2.00	462.31	1.5731	269.84	190.35	1446.3	0.7487	5.3142
4.00	497.26	1.5792	251.72	199.79	1448.4	0.7818	5.2886
6.00	534.32	1.5864	235.09	208.92	1450.2	0.8152	5.2632
10.00	614.68	1.6006	205.64	227.67	1454.0	0.8813	5.2138
15.00	728.03	1.6191	174.81	251.33	1458.4	0.9631	5.1538
20.00	856.87	1.6387	149.36	274.93	1462.2	1.0443	5.0955
25.00	1002.07	1.6586	128.28	298.96	1465.7	1.1245	5.0389
30.00	1165.55	1.6809	110.60	322.97	1468.6	1.2035	4.9837
35.00	1349.32	1.7000	95.775	347.24	1471.0	1.2820	4.9300
40.00	1553.84	1.7320	83.283	371.93	1472.8	1.3603	4.8774
45.00	1780.52	1.7672	72.592	396.60	1474.0	1.4373	4.8249
50.00	2031.70	1.7748	63.450	422.00	1474.4	1.5147	4.7728

THE PROPERTIES OF SUPERHEATED AMMONIA

T	P = 34.46kPa (Sat.Temp.=-52.84°C)			P = 48.24kPa (Sat.Temp.=-47.16°C)		
	v	h	s	v	h	s
-45.00	3195.1	1385.3	6.2959	2271.6	1382.6	6.1228
-40.00	3268.7	1396.1	6.3426	2325.4	1393.7	6.1709
-35.00	3342.3	1406.7	6.3879	2378.4	1404.6	6.2169
-30.00	3415.4	1417.2	6.4316	2430.7	1415.2	6.2620
-25.00	3488.8	1427.7	6.4753	2484.5	1426.3	6.3044
-20.00	3558.9	1438.1	6.5153	2535.8	1436.0	6.3490
-10.00	3705.0	1459.0	6.5975	2640.6	1457.8	6.4298
-5.00	3777.2	1469.6	6.6371	2692.2	1468.4	6.4696
0.00	3850.0	1480.0	6.6761	2744.0	1479.1	6.5086
5.00	3921.3	1490.5	6.7139	2795.5	1489.5	6.5469
10.00	3992.9	1501.2	6.7512	2847.3	1500.0	6.5846
15.00	4065.1	1511.7	6.7881	2889.3	1510.6	6.6215
20.00	4135.9	1521.9	6.8245	2949.9	1521.5	6.6578
25.00	4208.2	1532.9	6.8599	3012.0	1531.9	6.6940
30.00	4279.6	1543.4	6.8953	3053.1	1542.0	6.7283
35.00	4350.9	1553.1	6.9306	3104.2	1554.0	6.7656
40.00	4422.3	1565.7	6.9633	3155.6	1562.8	6.7978
50.00	4565.0	1586.0	7.0319	3257.7	1585.4	6.8663
55.00	4636.4	1596.8	7.0646	3308.6	1596.1	6.8992
60.00	4707.7	1607.4	7.0975	3359.9	1607.0	6.9321
65.00	4779.0	1618.3	7.1299	3411.0	1617.8	6.9645
70.00	4850.4	1629.2	7.1616	3461.6	1628.8	6.9963
75.00	4921.4	1640.1	7.1932	3512.5	1639.6	7.0279
80.00	4992.2	1651.1	7.2249	3564.0	1650.6	7.0595

T	P = 68.92kPa (Sat.Temp.=-40.74°C)			P = 96.49kPa (Sat.Temp.=-34.31°C)		
	v	h	s	v	h	s
-30.00	1693.5	1412.7	6.0784			
-25.00	1730.6	1423.6	6.1229	1228.3	1419.9	5.9474
-20.00	1767.8	1434.5	6.1662	1255.4	1431.4	5.9919
-15.00	1804.7	1445.2	6.2085	1282.1	1442.3	6.0354
-10.00	1842.0	1455.9	6.2499	1309.4	1453.2	6.0780
-5.00	1878.9	1466.6	6.2901	1336.0	1464.1	6.1185
0.00	1915.1	1477.3	6.3295	1362.5	1474.9	6.1584
5.00	1952.0	1487.9	6.3682	1388.8	1485.9	6.1979
10.00	1988.3	1498.6	6.4062	1415.2	1496.7	6.2367
15.00	2024.2	1509.3	6.4436	1441.5	1507.3	6.2745
20.00	2060.5	1519.8	6.4802	1468.0	1518.3	6.3111
25.00	2097.6	1531.1	6.5163	1492.9	1529.2	6.3481
30.00	2132.1	1540.6	6.5524	1519.9	1539.4	6.3843
35.00	2168.6	1553.1	6.5867	1545.7	1551.0	6.4185
40.00	2207.6	1561.8	6.6226	1571.8	1561.8	6.4543
50.00	2277.5	1584.4	6.6897	1623.3	1583.2	6.5222
55.00	2313.1	1595.2	6.7231	1649.0	1593.9	6.5556
60.00	2348.5	1606.1	6.7558	1674.9	1604.8	6.5888
65.00	2384.7	1616.8	6.7883	1700.9	1616.0	6.6217
70.00	2420.4	1628.2	6.8207	1726.6	1627.0	6.6538
75.00	2456.5	1638.9	6.8520	1751.6	1637.9	6.6854
80.00	2492.0	1650.0	6.8835	1777.9	1649.1	6.7168
85.00	2527.6	1661.0	6.9145	1803.2	1660.1	6.7481
90.00	2563.9	1672.2	6.9454	1828.7	1671.3	6.7793
100.00						

T	P = 124.05kPa (Sat.Temp.=-29.23°C)		
	v	h	s
-30.00			
-25.00	949.20	1416.4	5.8128
-20.00	970.69	1427.9	5.8591
-15.00	991.84	1439.4	5.9038
-10.00	1013.5	1450.5	5.9463
-5.00	1034.5	1461.6	5.9887
0.00	1055.4	1472.8	6.0295
5.00	1075.9	1483.8	6.0694
10.00	1096.9	1494.7	6.1085
15.00	1117.9	1505.6	6.1466
20.00	1137.6	1516.3	6.1842
25.00	1158.8	1527.5	6.2209
30.00	1179.0	1538.4	6.2565
35.00	1199.2	1549.6	6.2942
40.00	1219.7	1559.3	6.3259
50.00	1260.0	1581.4	6.3964
55.00	1280.0	1592.9	6.4300
60.00	1300.4	1604.0	6.4636
65.00	1320.7	1614.8	6.4966
70.00	1340.6	1626.0	6.5281
75.00	1360.3	1637.0	6.5607
80.00	1380.8	1648.1	6.5920
85.00	1400.6	1659.2	6.6233
90.00	1420.3	1670.3	6.6545
100.00	1462.0	1692.6	6.7151

T	P = 165.40kPa (Sat.Temp.=-23.10°C)			P = 206.75kPa (Sat.Temp.=-18.09°C)		
	v	h	s	v	h	s
-15.00	737.68	1434.7	5.7490	585.00	1430.0	5.6246
-10.00	753.63	1446.3	5.7937	598.60	1442.0	5.6717
-5.00	770.52	1457.8	5.8368	611.94	1454.0	5.7160
0.00	786.19	1469.3	5.8787	624.92	1465.5	5.7594
5.00	802.09	1480.5	5.9196	638.40	1477.1	5.8011
10.00	818.43	1491.6	5.9595	651.12	1488.6	5.8416
15.00	834.25	1502.7	5.9983	663.16	1499.8	5.8820
20.00	849.48	1514.0	6.0365	676.93	1511.4	5.9202
25.00	864.98	1524.6	6.0742	688.28	1522.1	5.8588
30.00	879.86	1536.1	6.1103	701.81	1533.9	5.9952
35.00	897.90	1547.5	6.1479	714.17	1544.2	6.0322
40.00	909.94	1557.1	6.1811	726.53	1556.8	6.0679
50.00	942.29	1580.1	6.2517	751.24	1578.1	6.1379
55.00	957.46	1591.0	6.2856	763.67	1589.4	6.1723
60.00	973.63	1602.1	6.3191	775.98	1600.5	6.2061
65.00	987.80	1613.1	6.3520	788.15	1611.5	6.2391
70.00	1003.0	1624.5	6.3844	800.90	1622.7	6.2726
75.00	1018.1	1635.1	6.4165	812.79	1633.7	6.3037
80.00	1033.7	1646.9	6.4492	824.25	1645.5	6.3370
90.00	1063.7	1669.0	6.5108	849.28	1667.6	6.3992
95.00	1078.7	1680.3	6.5416	861.09	1678.9	6.4303
105.00	1108.5	1702.9	6.6021	885.34	1701.5	6.4908
115.00	1139.0	1725.7	6.6617	909.51	1724.3	6.5503
125.00				933.39	1747.4	6.6091

T	P = 261.89kPa (Sat.Temp.=-12.54°C)		
	v	h	s
-10.00	467.30	1435.8	5.5374
-5.00	478.31	1448.5	5.5843
0.00	488.94	1460.5	5.6293
5.00	499.55	1472.5	5.6729
10.00	510.04	1484.4	5.7150
15.00	520.33	1496.0	5.7557
20.00	530.53	1507.4	5.7954
25.00	540.75	1519.0	5.8346
30.00	550.89	1530.3	5.8713
40.00	570.86	1552.9	5.9457
45.00	580.72	1564.2	5.9814
50.00	590.71	1575.5	6.0167
55.00	600.44	1586.9	6.0512
60.00	610.29	1598.1	6.0851
65.00	620.29	1609.4	6.1184
70.00	629.31	1620.7	6.1525
75.00	639.87	1632.1	6.1837
80.00	649.19	1642.8	6.2175
85.00	659.30	1654.8	6.2484
95.00	678.22	1677.3	6.3111
105.00	697.76	1700.1	6.3723
115.00	716.86	1723.9	6.4322
125.00	735.96	1745.9	6.4908
135.00	755.07	1769.1	6.5482
145.00			

h = kJ/kg, v = cm^3/g and s = kJ/kg·K

Datum - Sat. liquid h and s at -40°C.

THE PROPERTIES OF SUPERHEATED AMMONIA (continued)

T	P = 330.81kPa (Sat.Temp.=-6.78°C)		
	v	h	s
-10.00			
-5.00	373.66	1441.2	5.4488
0.00	382.49	1453.9	5.4965
5.00	391.20	1466.6	5.5420
10.00	399.79	1478.8	5.5856
15.00	408.22	1490.7	5.6280
20.00	416.56	1502.8	5.6686
25.00	424.74	1514.3	5.7092
30.00	433.06	1526.3	5.7462
40.00	449.16	1549.3	5.8223
45.00	457.10	1560.7	5.8592
50.00	465.05	1572.5	5.8948
55.00	472.95	1583.7	5.9299
60.00	480.82	1595.1	5.9645
65.00	488.68	1606.8	5.9984
70.00	496.50	1617.8	6.0320
75.00	504.29	1629.6	6.0652
80.00	5.1199	1640.8	6.0972
85.00	519.79	1652.1	6.1301
95.00	535.19	1675.2	6.1926
105.00	550.53	1698.0	6.2543
115.00	565.81	1721.0	6.3146
125.00	581.04	1744.3	6.3735
135.00	596.26	1767.7	6.4314
145.00	611.45	1791.2	6.4881

T	P = 413.51kPa (Sat.Temp.=-0.99°C)		
	v	h	s
5.00	308.77	1459.1	5.4117
10.00	315.89	1471.9	5.4575
15.00	322.86	1484.7	5.5014
20.00	329.75	1496.9	5.5450
25.00	336.69	1509.5	5.5849
30.00	343.20	1521.1	5.6265
40.00	356.54	1545.0	5.7031
45.00	363.08	1556.7	5.7402
50.00	369.57	1568.4	5.7767
55.00	376.04	1580.2	5.8122
60.00	382.43	1591.6	5.8473
65.00	388.77	1603.1	5.8820
70.00	395.17	1614.9	5.9160
75.00	401.41	1626.5	5.9496
80.00	407.72	1637.6	5.9819
90.00	420.22	1660.8	6.0469
95.00	426.47	1672.6	6.0784
100.00	432.69	1684.2	6.1096
105.00	438.87	1695.7	6.1404
115.00	451.17	1718.9	6.2011
125.00	463.48	1742.4	6.2604
135.00	475.74	1765.8	6.3186
145.00	487.94	1789.2	6.3759

T	P = 551.3kPa (Sat.Temp.=6.89°C)			P = 689.2kPa (Sat.Temp.=13.36°C)		
	v	h	s	v	h	s
10.00	231.73	1459.8	5.2833			
15.00	237.39	1473.5	5.3308	185.84	1461.9	5.1897
20.00	242.78	1487.1	5.3770	190.47	1476.0	5.2391
25.00	248.27	1499.4	5.4205	195.08	1489.8	5.2863
30.00	253.51	1512.8	5.4630	199.57	1503.6	5.3299
40.00	263.85	1537.4	5.5437	208.17	1529.5	5.4154
45.00	268.95	1549.7	5.5826	212.41	1542.1	5.4558
50.00	273.98	1561.7	5.6203	216.53	1554.7	5.4950
55.00	278.99	1573.7	5.6571	220.64	1567.2	5.5333
60.00	283.92	1585.6	5.6932	224.74	1579.5	5.5705
65.00	288.81	1597.5	5.7286	228.78	1591.7	5.6069
70.00	293.69	1609.5	5.7637	232.82	1603.7	5.6419
75.00	298.53	1621.3	5.7970	236.70	1616.0	5.6780
80.00	303.41	1632.5	5.8322	240.64	1628.0	5.7110
90.00	312.90	1656.3	5.8972	248.42	1651.8	5.7784
95.00	317.66	1668.2	5.9289	252.33	1663.7	5.8115
100.00	322.34	1680.0	5.9608	256.19	1675.7	5.8433
105.00	327.10	1691.8	5.9923	259.99	1687.6	5.8752
110.00	331.80	1703.5	6.0231	263.82	1699.5	5.9067
115.00	336.46	1715.2	6.0537	267.64	1711.5	5.9376
120.00	341.18	1727.1	6.0839	271.41	1723.4	5.9683
125.00	345.86	1738.9	6.1137	275.17	1735.3	5.9982
140.00	359.79	1774.5	6.2013	286.46	1771.4	6.0870
145.00	364.44	1786.7	6.2299	290.19	1783.5	6.1164

T	P = 964.9kPa (Sat.Temp.=23.77°C)			P = 1102.7kPa Sat.Temp.=28.13°C)			P = 1240.5kPa (Sat.Temp.=32.10°C)		
	v	h	s	v	h	s	v	h	s
30.00	137.57	1483.9	5.1165						
35.00	140.96	1498.6	5.1641	121.21	1489.0	5.0748	105.76	1479.3	4.9922
40.00	144.33	1512.6	5.2105	124.27	1504.1	5.1210	108.66	1495.2	5.0430
45.00	147.59	1526.9	5.2539	127.30	1518.8	5.1840	111.38	1510.3	5.0917
50.00	150.78	1540.4	5.2965	130.16	1532.8	5.1798	114.11	1525.2	5.1375
55.00	153.91	1553.6	5.3377	132.97	1546.7	5.2875	116.70	1539.5	5.1814
60.00	157.01	1566.7	5.3771	135.78	1560.2	5.2971	119.24	1553.5	5.2239
65.00	160.06	1579.7	5.4153	138.54	1573.4	5.2417	121.77	1567.3	5.2649
70.00	162.99	1592.5	5.4538	141.23	1586.5	5.3752	124.22	1580.8	5.3051
75.00	166.03	1605.3	5.4897	143.88	1599.9	5.4133	126.74	1594.0	5.3427
80.00	169.01	1617.3	5.5255	146.57	1612.3	5.4486	128.95	1667.1	5.3816
90.00	174.80	1642.6	5.5952	151.69	1637.9	5.5201	133.73	1633.2	5.4534
95.00	177.59	1655.2	5.6291	154.21	1650.4	5.5550	136.06	1645.8	5.4882
100.00	180.42	1667.2	5.6625	156.77	1663.1	5.5887	138.35	1658.1	5.5224
105.00	183.31	1679.6	5.6951	159.28	1675.6	5.6220	140.57	1673.8	5.5563
110.00	186.10	1691.7	5.7271	161.75	1687.9	5.6547	142.83	1684.0	5.5894
115.00	188.81	1704.0	5.7589	164.23	1700.3	5.6867	145.10	1682.6	5.6219
120.00	191.74	1716.2	5.7897	166.67	1712.6	5.7183	147.29	1739.2	5.6536
125.00	194.33	1728.8	5.8217	169.19	1725.0	5.7500	149.40	1677.9	5.6862
130.00	197.27	1740.4	5.8511	171.60	1737.3	5.7797	152.16	1734.4	5.7162
140.00	202.65	1765.2	5.9115	176.43	1762.0	5.8409	156.08	1758.9	5.7776
145.00	205.35	1777.3	5.9407	178.85	1774.4	5.8705	158.20	1771.4	5.8074
150.00	208.05	1789.5	5.9697	181.26	1786.7	5.8998	160.41	1783.7	5.8371
160.00	213.5	1814.0	6.0269	186.03	1811.4	5.9574	166.87	1808.6	5.8954
170.00				190.75	1836.1	6.0139	170.29	1833.7	5.9518

T	P = 1378.4kPa (Sat.Temp.=35.74°C)		
	v	h	s
45.00	98.650	1501.5	5.0186
50.00	101.22	1517.2	5.0669
55.00	103.63	1532.1	5.1129
60.00	106.00	1546.5	5.1569
65.00	108.38	1560.8	5.1991
70.00	110.63	1574.5	5.2401
75.00	112.82	1588.7	5.2800
80.00	115.17	1601.4	5.3171
85.00	117.18	1615.2	5.3568
95.00	121.50	1641.3	5.4276
100.00	123.56	1654.3	5.4626
105.00	125.65	1667.3	5.4969
110.00	127.73	1680.0	5.5303
115.00	129.77	1692.5	5.5632
120.00	131.71	1705.4	5.5958
125.00	133.89	1717.9	5.6272
130.00	135.70	1730.5	5.6601
135.00	137.82	1743.0	5.6894
145.00	141.74	1768.1	5.7509
150.00	143.71	1780.7	5.7806
160.00	147.58	1805.8	5.8393
170.00	151.40	1830.9	5.8967
180.00	155.22	1856.1	5.9527
190.00	159.05	1881.2	6.0075

T	P = 1516.2kPa (Sat.Temp=39.12°C)			P = 1654.0kPa (Sat.Temp.=42.27°C)			P = 1791.9kPa (Sat.Temp.=45.23°C)			P = 2067.5kPa (Sat.Tem.=50.67°C)		
	v	h	s	v	h	s	v	h	s	v	h	s
45.00	88.195	1492.6	4.9502	79.492	1483.3	4.8840						
50.00	90.633	1508.9	4.9999	81.778	1500.3	4.9368	74.257	1491.2	4.8759			
55.00	92.947	1524.5	5.0486	84.032	1516.4	4.9869	76.403	1508.4	4.9280	64.097	1490.6	4.8157
60.00	95.202	1539.5	5.0941	86.150	1532.1	5.0346	78.472	1524.7	4.9777	66.049	1508.6	4.8701
65.00	97.393	1554.2	5.1369	88.168	1547.5	5.0800	80.440	1540.4	5.0249	67.878	1525.8	4.9211
70.00	99.536	1568.6	5.1798	90.216	1562.0	5.1234	82.372	1555.7	5.0701	69.783	1542.5	4.9694
75.00	101.50	1582.3	5.2209	92.130	1576.7	5.1643	84.211	1570.7	5.1118	71.352	1557.6	5.0157
80.00	103.65	1596.7	5.2598	94.198	1590.9	5.2066	86.030	1584.9	5.1557	73.120	1573.9	5.0579
85.00	105.77	1610.1	5.2978	95.923	1604.3	5.2435	87.841	1600.3	5.1928	74.763	1588.6	5.1028
95.00	109.61	1636.8	5.3714	99.680	1631.9	5.3198	91.289	1627.3	5.2709	77.873	1617.7	5.1813
100.00	111.53	1649.9	5.4073	101.46	1645.3	5.3557	93.029	1640.8	5.3078	79.443	1631.6	5.2198
105.00	113.46	1663.8	5.4420	103.28	1658.7	5.3910	94.696	1654.3	5.3437	80.948	1645.5	5.2567
110.00	115.37	1675.8	5.4759	105.07	1671.9	5.4257	96.326	1667.7	5.3788	82.405	1659.3	5.2930
115.00	117.21	1688.8	5.5095	106.80	1684.9	5.4598	97.967	1680.9	5.4133	83.856	1672.9	5.3286
120.00	119.10	1701.8	5.5423	108.56	1697.6	5.4923	99.568	1693.9	5.4460	85.342	1686.5	5.3626
125.00	120.89	1714.0	5.5748	110.25	1711.1	5.5263	101.21	1707.5	5.4812	86.723	1700.0	5.3968
130.00	122.81	1727.6	5.6060	111.93	1723.3	5.5577	102.86	1719.86	5.5112	88.196	1713.2	5.4307
135.00	124.60	1739.4	5.6380	113.63	1736.7	5.5891	104.30	1733.8	5.5465	89.496	1726.5	5.4616
145.00	128.20	1765.1	5.6991	116.99	1762.2	5.6514	107.50	1759.0	5.6069	92.345	1752.8	5.5264
150.00	130.00	1777.7	5.7292	118.67	1774.9	5.6815	109.03	1771.9	5.6376	93.691	1765.9	5.5573
160.00	133.60	1803.0	5.7883	121.98	1800.2	5.7409	112.12	1797.4	5.6974	96.389	1791.9	5.6183
170.00	137.14	1828.4	5.8460	125.25	1825.8	5.7990	115.16	1824.0	5.7559	99.087	1817.8	5.6775
180.00	140.63	1853.7	5.9023	128.46	1851.2	5.8558	118.19	1848.8	5.8130	101.74	1843.8	5.7354
190.00	144.10	1879.0	5.9576	131.64	1876.4	5.9114	121.2	1874.2	5.8689	104.36	1869.7	5.7920

h = kJ/kg, v = cm^3/g and s = kJ/kg·K

T = ^{o}C. P = kPa

THE PROPERTIES OF SATURATED,CARBON DIOXIDE

T	P	v_f	v_g	h_f	h_g	s_f	s_g	T	P	v_f	v_g	h_f	h_g	s_f	s_g
-95.00	23 42	0.6298	1444.6	-282.22	300.71	2.5426	5.8260	-30.00	1426.7	0.9308	27.002	18.76	322.28	3.9394	5.1878
-85.00	59.83	0.6393	602.03	-276.55	306.48	2.6091	5.7275	-25.00	1662.0	0.9503	22.884	29.52	322.26	3.9792	5.1587
-75.00	133.70	0.6412	268.80	-255.91	311.74	2.6683	5.5466	-20.00	1967.2	0.9708	19.470	38.07	322.33	4.0122	5.1328
-70.00	198.35	0.6479	185.64	-252.44	313.47	2.7049	5.4844	-15.00	2289.4	0.9906	16.599	50.82	324.25	4.0662	5.1246
-65.00	286.98	0.6511	128.85	-242.60	314.98	2.7425	5.4235	-5.00	3044.4	1.0478	12.144	72.56	321.53	4.1426	5.0713
-60.00	410.05	0.6576	91.451	-235.69	315.89	2.7864	5.3694	0.00	3483.9	1.0817	10.375	85.10	319.87	4.1864	5.0451
-56.61	517.57	0.8490	72.229	-31.86	316.05	3.7200	5.3273	5.00	3970.1	1.1321	8.847	99.37	317.11	4.2323	5.0168
-50.00	682.63	0.8667	55.411	-19.49	317.88	3.7767	5.2883	10.00	4504.5	1.1662	7.5225	112.56	313.95	4.2781	4.9894
-45.00	632.56	0.8815	45.813	-9.79	319.25	3.8187	5.2610	15.00	5090.7	1.2203	6.3318	127.16	308.13	4.3284	4.9550
-40.00	1004.2	0.8971	38.162	0.00	320.47	3.8594	5.2348	20.00	5731.7	1.3032	5.2465	144.45	298.89	4.3840	4.9171
-35.00	1201.8	0.9133	32.001	9.48	321.21	3.8996	5.2080	25.00	6428.7	1.3964	4.1991	162.56	284.83	4.4457	4.8688
								30.00	7202.0	1.9099	2.6387	207.36	242.38	4.5908	4.7116

THE PROPERTIES OF SUPERHEATED CARBON DIOXIDE

T	P = 6.89kPa			P = 68.92kPa			P = 137.84kPa			P = 275.67kPa			P = 551.34kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
-55.00	5983.8	662.13	6.2013	591.68	660.86	5.7662	292.31	659.45	5.6341	143.00	656.63	5.4967			
-45.00	6258.0	670.24	6.2369	619.74	669.07	5.8019	307.20	667.92	5.6700	150.49	665.35	5.5331	72.273	660.48	5.3521
-15.00	7080.6	694.75	6.3359	703.78	693.61	5.9009	348.89	692.42	5.7695	173.10	690.20	5.6340	84.524	685.55	5.4723
10.00	7766.0	715.58	6.4125	772.85	714.65	5.9775	382.00	713.49	5.8464	190.59	711.40	5.7116	93.517	707.21	5.5617
35.00	8451.4	737.15	6.4843	841.87	736.44	6.0493	419.03	735.51	5.9183	207.74	733.52	5.7842	103.23	730.19	5.6397
65.00	9274.1	763.23	6.5651	924.91	762.76	6.1301	461.65	762.05	5.9991	229.86	760.89	5.8659	113.93	758.53	5.7256
90.00	9958.8	785.55	6.6286	993.42	785.31	6.1936	495.77	784.87	6.0625	246.91	783.88	5.9302	122.52	782.31	5.7922
150.00	11604.	841.52	6.7704	1158.6	841.28	6.3354	578.79	840.83	6.2044	288.86	840.36	6.0729	143.85	838.75	5.9381
205.00	13106.	895.21	6.8889	1310.0	894.98	6.4539	654.39	894.75	6.3228	326.90	894.29	6.1922	163.00	893.36	6.0586
315.00	16122.	1009.6	7.1036	1612.2	1009.6	6.6686	805.81	1009.4	6.5375	402.81	1009.2	6.4069	201.20	1008.7	6.2754
425.00	19132.	1130.9	7.2920	1913.9	1130.9	6.8569	956.60	1130.9	6.7259	477.43	1130.6	6.5953	239.10	1130.4	6.4642
535.00	22148.	1258.1	7.4606	2215.4	1258.1	7.0256	1107.4	1258.1	6.8946	553.81	1258.1	6.7639	276.72	1257.9	6.6329
650.00	25301.	1395.8	7.6202	2530.7	1395.8	7.1852	1265.1	1395.8	7.0541	632.55	1395.8	6.9235	316.27	1395.5	6.7925
760.00	28317.	1531.6	7.7594	2831.7	1531.6	7.3244	1415.9	1531.6	7.1933	709.80	1531.6	7.0627	353.97	1531.6	6.9317
870.00	31327.	1670.2	7.8859	3133.3	1670.2	7.4509	1566.0	1670.2	7.3198	783.37	1670.2	7.1892	391.73	1670.2	7.0581
980.00	34336.	1811.1	8.0040	3434.4	1811.1	7.5690	1716.8	1811.1	7.4380	858.55	1811.1	7.3074	429.20	1811.1	7.1763

T	P = 827.02kPa			P = 1102.6kPa			P = 1378.4kPa			P = 1654.0kPa			P = 2067.5kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
-15.00	54.975	680.71	5.3848	40.121	676.15	5.3146	31.093	671.59	5.2548	25.152	666.75	5.2030			
10.00	61.173	702.79	5.4788	44.992	698.60	5.4127	35.284	694.42	5.3612	28.804	690.00	5.3147	22.243	683.72	5.2595
35.00	67.032	726.56	5.5588	49.406	722.94	5.4986	39.227	719.32	5.4497	32.177	715.69	5.4084	25.170	710.42	5.3546
65.00	75.289	756.19	5.6456	56.006	753.84	5.5878	44.407	751.49	5.5413	36.685	749.14	5.5020	28.892	745.62	5.4533
90.00	81.024	780.64	5.7128	60.300	779.02	5.6555	47.852	777.40	5.6101	39.554	775.74	5.5713	31.267	773.41	5.5253
150.00	95.465	837.59	5.8611	71.303	836.21	5.8042	56.813	834.82	5.7607	47.152	833.67	5.7263	37.476	831.58	5.6800
205.00	108.38	892.43	5.9820	81.068	891.51	5.9259	64.693	890.59	5.8832	53.786	889.65	5.8472	42.930	888.03	5.8049
315.00	133.97	1008.3	6.1988	100.41	1007.8	6.1431	80.267	1007.3	6.1008	66.802	1006.6	6.0657	53.363	1005.[illegible]	6.0234
425.00	159.37	1130.2	6.3876	119.45	1129.7	6.3332	95.534	1129.5	6.2913	79.272	1129.3	6.2557	63.580	1128.8	6.2138
535.00	184.53	1257.9	6.5563	138.36	1257.7	6.5018	110.68	1257.7	6.4600	92.196	1257.4	6.4252	73.785	1257.2	6.3833
650.00	210.82	1395.5	6.7158	158.13	1395.5	6.6614	126.51	1395.5	6.6195	105.44	1395.5	6.5848	84.317	1395.3	6.5429
760.00	236.04	1531.6	6.8550	177.05	1531.6	6.8006	141.65	1531.4	6.7588	118.05	1531.4	6.7240	94.453	1531.4	6.6821
870.00	261.19	1670.2	6.9815	195.90	1670.2	6.9271	156.73	1670.2	6.8852	130.66	1670.2	6.8505	104.53	1670.0	6.8086
980.00	286.17	1811.1	7.0997	214.68	1811.1	7.0453	171.81	1811.1	7.0034	143.15	1811.1	6.9686	114.54	1811.1	6.9268

T	P = 2481.1kPa			P = 3032.4kPa			P = 3583.7kPa			P = 4135.1kPa			P = 5513.4kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
10.00	17.842	677.21	5.2067	13.834	668.84	5.1422	11.062	660.23	5.0861	9.0645	651.86	5.0325			
35.00	20.544	704.81	5.3065	16.255	697.57	5.2473	13.313	690.60	5.1962	11.163	683.36	5.1475	7.2510	665.24	5.0576
65.00	23.551	742.10	5.4097	18.941	737.40	5.3560	15.655	732.71	5.3099	13.222	728.01	5.2663	9.2286	716.27	5.1857
90.00	25.706	770.79	5.4851	20.695	767.56	5.4352	17.213	764.26	5.3926	14.666	761.04	5.3520	10.514	752.81	5.2768
150.00	31.032	829.73	5.6444	25.185	826.96	5.6005	21.130	824.42	5.5633	18.150	821.66	5.5287	13.318	815.18	5.4631
205.00	35.600	886.64	5.7702	28.962	884.79	5.7292	24.385	882.93	5.6949	21.023	881.07	5.6639	15.561	876.21	5.6032
315.00	44.985	1005.2	5.9890	36.279	1004.3	5.9505	30.633	1003.3	5.9183	26.505	1001.9	5.8902	19.788	999.61	5.8332
425.00	52.950	1128.3	6.1795	43.279	1127.8	6.1414	36.625	1127.2	6.1095	31.716	1126.7	6.0824	23.739	1125.3	6.0279
535.00	61.425	1256.9	6.3490	50.261	1256.5	6.3109	42.502	1256.0	6.2791	36.835	1255.8	6.2519	27.596	1254.8	6.1975
650.00	70.317	1395.3	6.5086	57.510	1395.1	6.4705	48.660	1394.9	6.4387	42.172	1394.4	6.4115	31.627	1393.9	6.3570
760.00	78.722	1531.4	6.6478	64.426	1531.2	6.6097	54.518	1530.9	6.5779	47.264	1530.7	6.5507	35.459	1530.2	6.4962
870.00	87.127	1670.0	6.7743	71.224	1669.8	6.7362	60.322	1669.8	6.7019	52.295	1669.8	6.6771	39.242	1669.3	6.6227
980.00	95.534	1810.8	6.8924	78.198	1810.8	6.8543	66.178	1810.9	6.8223	57.342	1810.6	6.7953	43.035	1810.4	6.7409

T	P = 6891.8kPa			P = 8270.1kPa			P = 9648.5kPa			P = 12405kPa			P = 11026kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
65.00	6.8454	704.53	5.1132												
90.00	8.0288	744.66	5.2097	6.3717	737.30	5.1552									
150.00	10.421	808.48	5.4055	8.4998	801.79	5.3587	7.1234	795.31	5.3220						
205.00	12.291	871.58	5.5522	10.135	866.94	5.5104	8.5983	862.07	5.4769	6.5474	847.28	5.4150	7.4461	855.60	5.4451
315.00	15.752	997.28	5.7884	13.071	994.95	5.7516	11.151	992.38	5.7214	8.5867	987.48	5.6674	9.7078	990.05	5.6938
425.00	18.982	1123.6	5.9852	15.761	1122.5	5.9492	13.452	1121.1	5.9194	10.430	1118.2	5.8683	11.821	1119.6	5.8937
535.00	22.065	1253.9	6.1556	18.369	1253.0	6.1204	15.730	1252.0	6.0911	12.216	1250.4	6.0418	13.753	1251.3	6.0654
650.00	25.308	1393.2	6.3152	21.089	1392.5	6.2800	18.076	1392.1	6.2511	14.058	1390.7	6.2030	15.814	1391.4	6.2259
760.00	28.373	1529.8	6.4544	23.654	1529.5	6.4192	20.283	1529.1	6.3903	15.782	1527.9	6.3422	17.748	1528.4	6.3610
870.00	31.414	1669.1	6.5808	26.188	1668.8	6.5457	22.460	1668.4	6.5168	17.488	1667.9	6.4686	19.664	1668.1	6.4916
980.00	34.437	1810.2	6.6990	28.722	1809.9	6.6638	24.025	1809.9	6.6350	19.187	1809.5	6.5868	21.568	1809.7	6.6102

Datum - Sat. liquid h = 0 at -40^{o}C.

h = kJ/kg, v = cm^3/g and s = kJ/kg·K

Below a triple point temperature of -56.6^{o}C the values shown are for solid and gaseous phases only

T = °C. P = kPa

THE PROPERTIES OF SATURATED ETHYLENE

T	P	v_f	v_g	h_f	h_g	s_f	s_g
-165.00	.23		135188.	-146.32	415.54	-1.0480	4.1683
-180.00	.51		66007.	-136.03	421.43	-0.9509	3.9762
-150.00	2.08		18236.	-111.60	432.72	-0.7678	3.6544
-140.00	6.14		6436.3	-87.35	445.12	-0.5786	3.4206
-130.00	15.61		2697.5	-83.42	462.69	-0.4113	3.2440
-120.00	34.44		1299.8	-38.74	458.92	-0.2225	3.0578
-110.00	69.13		682.32	-15.69	485.10	-0.1083	2.9446
-105.00	93.50	1.7420	518.14	-2.90	482.78	-0.0082	2.8797
-95.00	164.28	1.7905	308.77	21.48	490.06	0.1161	2.7459
-85.00	268.28	1.8382	199.36	46.98	497.78	0.2672	2.6627
-75.00	441.31	1.9511	98.020	72.22	504.26	0.3894	2.5701
-60.00	775.02	1.9994	90.801	111.17	512.61	0.6044	2.4879
-50.00	1048.0	2.0749	51.537	137.64	516.28	0.7074	2.4044
-40.00	1422.0	2.1650	37.020	164.65	518.14	0.8101	2.3262
-30.00	1927.8	2.2914	26.996	193.68	516.76	0.9268	2.2564
-20.00	2528.0	2.3849	20.142	225.46	516.67	1.0497	2.1991
-5.00	3639.0	2.7003	12.580	286.18	503.30	1.2714	2.0811
5.00	4570.5	3.2157	8.3607	350.43	474.99	1.4949	1.9427
10.00	5128.9	4.4060	4.2584	400.72	397.13	1.6636	1.6508

THE PROPERTIES OF SUPERHEATED ETHYLENE

T	P = 101.3kPa			P = 202.7kPa			P = 405.2kPa			P = 607.8kPa		
	v	h	s	v	h	s	v	h	s	v	h	s
-95.00	504.97	494.26	2.9135									
-85.00	536.21	507.03	2.9851	260.66	501.41	2.7543						
-70.00	582.59	526.02	3.0854	284.97	521.67	2.8532	135.97	512.73	2.6324			
-60.00	613.38	539.24	3.1443	301.03	535.30	2.9157	144.65	527.65	2.6977	92.551	518.75	2.5577
-50.00	642.98	552.39	3.2058	316.82	549.01	2.9797	153.24	542.06	2.7616	98.353	534.57	2.6233
-40.00	673.85	565.81	3.2657	332.37	562.79	3.0396	161.56	556.51	2.8261	104.44	550.23	2.6879
-25.00	713.89	586.59	3.3496	355.56	583.95	3.1235	173.84	578.45	2.9136	115.70	573.03	2.7816
-15.00	754.00	600.70	3.4 78	370.88	598.52	3.1816	181.88	593.59	2.9727	111.35	588.59	2.8437
-5.00	783.36	614.92	3.4629	386.09	612.79	3.2370	189.79	608.48	3.0330	120.56	603.90	2.9025
5.00	808.47	629.69	3.5156	401.25	627.61	3.2894	197.64	623.66	3.0884	131.19	619.59	2.9590
15.00	838.18	644.97	3.5686	416.36	643.08	3.3425	205.39	639.36	3.1412	135.05	635.59	3.0154
35.00	897.64	676.08	3.6717	446.46	674.54	3.4480	220.82	671.41	3.2494	145.63	668.35	3.1249
60.00	971.38	716.98	3.8018	483.75	715.35	3.5797	239.91	713.02	3.3787	158.63	710.23	3.2531
80.00	1030.4	751.45	3.8974	513.73	750.12	3.6741	255.10	748.06	3.4757	168.91	745.58	3.3529
105.00	1104.1	796.36	4.0259	550.72	795.23	3.8081	273.85	793.14	3.6033	181.63	791.27	3.4816
125.00	1163.1	833.99	4.1200	580.63	833.29	3.9020	288.85	831.42	3.7013	191.89	629.53	3.5756
160.00	1266.0	903.26	4.2915	631.77	902.56	4.0779	315.01	900.93	3.8726	209.26	899.53	3.7514
195.00	1369.0	976.63	4.4540	683.56	976.18	4.2405	341.03	974.80	4.0397	226.82	973.40	3.9182
225.00	1457.2	1043.6	4.5896	727.96	1043.2	4.3799	363.22	1042.0	4.1792	241.71	1040.8	4.0577
200.00	1560.1	1123.02	4.7436	779.10	1122.6	4.5343	389.17	1121.6	4.3333	259.08	1120.5	4.2119

T	P = 1013kPa			P = 1520kPa			P = 2026kPa			P = 3039kPa		
	v	h	s	v	h	s	v	h	s	v	h	s
-50.00	54.194	519.22	2.4298									
-40.00	58.432	536.74	2.5079									
-25.00	64.263	561.80	2.6010	39.600	544.43	2.4293						
-15.00	68.135	577.93	2.6639	42.551	563.02	2.5023	29.335	546.57	2.3920			
-5.00	71.736	594.42	2.7281	45.262	581.42	2.5746	31.814	567.39	2.4301	17.958	532.32	2.2263
5.00	75.239	611.17	2.7872	47.904	599.88	2.6411	34.148	587.36	2.4817	19.960	558.34	2.3159
15.00	78.726	627.90	2.8437	50.487	618.09	2.7020	36.288	606.58	2.5845	21.766	581.93	2.3988
35.00	85.359	661.65	2.9595	55.354	653.30	2.8131	40.157	644.44	2.7032	24.809	625.44	2.5459
60.00	93.579	704.88	3.0940	61.054	697.91	2.9601	44.761	690.47	2.8554	28.342	674.88	2.7047
80.00	99.990	740.90	3.1938	65.484	734.75	3.0569	48.196	728.66	2.9570	30.796	715.34	2.8167
105.00	107.79	787.09	3.3269	70.909	782.00	3.1971	52.409	776.68	3.1007	33.967	765.54	2.9588
125.00	103.97	825.78	3.4248	75.122	821.32	3.2956	55.726	816.63	3.1984	36.330	806.53	3.0644
160.00	124.79	896.28	3.6006	82.530	892.56	3.4709	61.367	888.60	3.3787	40.266	880.47	3.2448
195.00	135.41	970.64	3.7674	89.802	967.40	3.6377	66.936	964.17	3.5456	44.129	957.49	3.4160
225.00	144.51	1038.5	3.9031	95.925	1035.7	3.7733	71.599	1032.6	3.6850	47.274	1027.0	3.5554
260.00	155.01	1118.4	4.0570	103.01	1116.1	3.9272	76.974	1113.5	3.8393	51.004	1108.6	3.7137

T	P = 4052kPa			P = 50.65kPa			P = 6078kPa			P = 8104kPa		
	v	h	s	v	h	s	v	h	s	v	h	s
5.00	12.120	516.28	2.1133									
15.00	14.159	550.16	2.2302	9.4991	523.84	2.0998	5.5428	444.63	1.7917	3.3873	369.09	1.6150
35.00	16.959	603.05	2.4061	12.008	573.53	2.2794	8.5626	545.78	2.1329	4.7435	459.65	1.9794
60.00	20.164	659.07	2.5833	15.170	640.93	2.4702	11.611	621.88	2.3781	7.5538	577.44	2.1855
80.00	22.184	701.63	2.7066	16.860	686.87	2.8049	13.472	671.94	2.5263	9.1955	639.48	2.3830
105.00	24.710	754.19	2.8541	19.323	742.38	2.7621	15.525	730.60	2.6826	10.896	705.61	2.5491
125.00	26.565	798.67	2.9596	20.723	786.33	2.8714	16.933	776.26	2.7995	12.077	755.07	2.6735
160.00	29.716	872.33	3.1443	23.410	863.95	3.0606	19.165	855.35	2.9977	13.984	838.37	2.8805
195.00	32.662	950.80	3.3196	25.498	932.26	3.2156	21.024	925.55	3.1490	15.450	911.10	3.0434
225.00	35.134	1021.2	3.4591	27.853	1015.1	3.3836	23.067	1009.1	3.3202	17.089	997.73	3.2159
260.00	38.019	1103.5	3.6216	30.215	1098.6	3.5462	25.096	1093.7	3.4876	18.791	1084.0	3.3829

T	P = 10 130kPa			P = 15 195kPa			P = 20 260kPa			P = 25 325kPa			P = 30 390kPa		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
20.00	2.8209	345.44	1.4120	2.4768	337.25	1.3345	2.3893	333.90	1.2704	2.2657	331.97	1.2333	2.1977	331.25	1.1968
35.00	3.4254	408.79	1.6241	2.7603	382.83	1.4845	2.5910	373.84	1.4029	2.4385	368.96	1.3599	2.3682	368.24	1.3209
60.00	5.1191	530.93	2.0097	3.4960	464.65	1.7459	2.9341	444.42	1.6412	2.6844	436.05	1.5742	2.5595	433.26	1.5240
80.00	6.6400	604.69	2.2162	4.0985	538.74	1.9590	3.3034	507.90	1.8213	3.0150	493.54	1.7354	2.7963	487.09	1.6712
105.00	8.2067	679.03	2.3276	5.0804	623.24	2.1982	3.8781	588.54	2.0437	3.3753	568.89	1.9385	3.0879	556.65	1.8711
125.00	9.2195	733.13	2.5594	5.8136	684.71	2.3492	4.3235	650.17	2.2023	3.6532	628.36	2.0931	3.3495	613.54	2.0142
160.00	10.803	821.63	2.7842	6.9295	782.56	2.5916	5.3064	752.09	2.4493	4.3075	736.47	2.3488	3.8081	715.58	2.2651
190.00	12.158	897.56	2.9508	7.8882	865.25	2.7773	5.9874	838.60	2.6424	4.8135	817.76	2.5420	4.2556	802.86	2.4621
225.00	13.005	986.49	3.1355	8.8834	960.22	2.9722	6.7063	937.04	2.8463	5.4590	918.73	2.7495	4.7840	904.91	2.6701
260.00	14.920	1074.4	3.3034	9.9260	1052.6	3.1485	7.4289	1033.0	3.0312	6.1804	1017.4	2.9391	5.1815	1004.7	2.8638

h = kJ/kg, v = cm^3/g and s = kJ/kg·K

Datum - Sat. liquid h and s = 0
at 1 atmosphere pressure

T = °C. P = kPa

THE PROPERTIES OF SATURATED FREON-21

Datum - Sat. liquid h and s = 0 at -40°C.

T	P	v_f	v_g	h_f	h_g	s_f	s_g
-40.00	9.36	0.6605	2003.3	0.00	266.42	0.0033	1.1430
-35.00	12.60	0.6646	1519.0	4.95	268.94	0.2090	1.1295
-30.00	16.70	0.6715	1166.5	9.80	271.43	0.0403	1.1182
-25.00	21.91	0.6726	908.48	14.94	273.99	0.0629	1.1053
-20.00	28.25	0.6816	716.33	19.63	276.47	0.0796	1.0957
-15.00	36.13	0.6848	570.10	24.72	279.00	0.1005	1.0856
-10.00	45.67	0.6905	458.76	29.71	281.54	0.1197	1.0770
-5.00	57.08	0.6960	372.79	34.70	284.03	0.1385	1.0685
0.00	70.75	0.7018	305.53	39.76	286.58	0.1568	1.0615
5.00	86.80	0.7072	252.53	44.84	289.08	0.1755	1.0533
10.00	105.65	0.7129	210.38	49.98	291.60	0.1938	1.0475
15.00	127.76	0.7195	176.47	55.21	294.15	0.2116	1.0438
20.00	152.64	0.7244	148.93	60.39	296.65	0.2310	1.0340
25.00	183.38	0.7326	126.61	65.76	299.11	0.2469	1.0323
30.00	215.23	0.7385	108.19	71.07	301.58	0.2654	1.0258
35.00	252.98	0.7452	92.889	76.50	304.00	0.2829	1.0213
40.00	294.99	0.7525	80.289	81.86	306.38	0.3004	1.0175
45.00	345.87	0.7595	69.598	87.45	308.74	0.3176	1.0135
50.00	395.80	0.7671	60.632	93.00	311.04	0.3351	1.0099
55.00	456.26	0.7750	53.046	98.61	313.29	0.3523	1.0065
60.00	521.85	0.7828	46.552	104.33	315.49	0.3693	1.0032
65.00	582.37	0.7912	40.994	110.10	317.63	0.3862	1.0001
70.00	663.95	0.7998	36.223	115.95	319.74	0.4032	0.9975

THE PROPERTIES OF SUPERHEATED FREON-21

T	P = 8.27kPa (Sat.Temp.=-42.00°C)		
	v	h	s
-40.00	2268.6	266.44	1.1530
-35.00	2314.8	269.02	1.1640
-30.00	2375.6	271.60	1.1748
-25.00	2401.2	274.18	1.1857
-20.00	2479.4	276.83	1.1959
-10.00	2562.8	282.11	1.2162
-5.00	2611.1	284.81	1.2263
0.00	2660.5	287.54	1.2366
5.00	2709.8	290.28	1.2467
10.00	2758.7	293.05	1.2565
15.00	2807.6	295.85	1.2663
20.00	2856.6	298.69	1.2759
25.00	2904.6	301.57	1.2860
30.00	2955.4	304.47	1.2950
35.00	3002.0	307.41	1.3060
40.00	3052.6	310.31	1.3121
50.00	3150.4	316.27	1.3328
55.00	3199.2	319.27	1.3421
60.00	3248.1	322.30	1.3515
65.00	3297.0	325.38	1.3606
70.00	3345.9	328.49	1.3696
75.00	3394.8	331.63	1.3786
80.00	3443.6	334.78	1.3877
85.00	3492.5	337.95	1.3967
90.00	3541.4	341.14	1.4057
100.00	3639.2	347.58	1.4205
105.00	3688.0	350.83	1.4231
110.00	3736.9	354.09	1.4185
115.00	3785.8	357.38	1.4020
120.00	3834.7	360.69	1.3675

T	P = 13.78kPa (Sat.Temp.=-33.44°C)		
	v	h	s
-30.00	1416.1	271.47	1.1333
-25.00	1447.8	274.13	1.1434
-15.00	1505.8	279.36	1.1645
-10.00	1534.7	282.02	1.1747
-5.00	1564.4	284.72	1.1852
0.00	1594.2	287.45	1.1942
5.00	1623.1	290.19	1.2061
10.00	1652.5	292.98	1.2150
15.00	1682.5	295.81	1.2220
20.00	1711.4	298.62	1.2350
25.00	1740.6	301.54	1.2438
30.00	1711.7	304.42	1.2536
40.00	1829.3	310.27	1.2726
45.00	1859.1	313.22	1.2825
50.00	1887.9	316.20	1.2914
55.00	1917.4	319.23	1.3007
60.00	1947.1	322.26	1.3100
65.00	1976.5	325.31	1.3191
70.00	2005.4	328.43	1.3280
75.00	2034.7	331.54	1.3377
80.00	2065.8	334.74	1.3451
85.00	2092.6	337.83	1.3567
90.00	2124.8	341.03	1.3639
100.00	2182.1	347.56	1.3819
105.00	2211.3	350.86	1.3906
110.00	2240.5	354.16	1.3992
115.00	2270.3	357.49	1.4079
120.00	2299.8	360.99	1.4153

T	P = 27.57kPa (Sat.Temp.=-20.50°C)		
	v	h	s
-15.00	749.51	279.12	1.1080
-10.00	764.60	281.81	1.1183
-5.00	778.70	284.50	1.1283
0.00	793.74	287.25	1.1386
5.00	808.93	290.00	1.1487
10.00	823.42	292.79	1.1585
15.00	837.96	295.63	1.1684
20.00	852.85	298.44	1.1779
25.00	868.68	301.37	1.1880
30.00	881.59	304.17	1.1975
35.00	898.50	307.22	1.2063
40.00	910.07	310.00	1.2179
50.00	941.32	316.03	1.2353
55.00	956.22	319.03	1.2444
60.00	971.38	322.07	1.2535
65.00	986.18	325.14	1.2629
70.00	1000.3	328.26	1.2723
75.00	1015.8	331.39	1.2804
80.00	1031.0	334.54	1.2911
90.00	1059.9	340.91	1.3079
95.00	1074.0	344.16	1.3167
100.00	1089.1	347.42	1.3254
105.00	1104.1	350.72	1.3340
110.00	1118.7	354.02	1.3427
115.00	1133.3	357.35	1.3514
120.00	1147.9	360.71	1.3600
125.00	1162.5	364.10	1.3687
130.00	1177.4	367.50	1.3772

T	P = 68.92kPa Sat.Temp.=-0.61°C)		
	v	h	s
5.00	19.97	289.37	1.0733
10.00	325.81	292.16	1.0831
15.00	331.49	295.01	1.0929
20.00	338.75	297.86	1.1029
25.00	343.87	300.80	1.1123
30.00	349.86	303.60	1.1226
40.00	361.84	309.53	1.1414
45.00	367.80	312.51	1.1506
50.00	373.79	315.50	1.1604
55.00	379.71	318.53	1.1694
60.00	385.68	321.58	1.1786
65.00	391.70	324.66	1.1881
70.00	397.65	327.78	1.1970
75.00	403.61	330.90	1.2067
80.00	409.56	334.09	1.2141
85.00	415.52	337.32	1.2256
90.00	421.50	340.41	1.2322
100.00	433.39	346.98	1.2508
105.00	439.33	350.28	1.2595
110.00	445.30	353.60	1.2682
115.00	452.24	356.95	1.2770
120.00	451.10	360.33	1.2851
125.00	478.38	363.68	1.2948
130.00	464.53	367.13	1.3024
135.00	480.07	370.56	1.3111
140.00	485.27	374.04	1.3196
150.00	494.08	381.06	1.3362
155.00	496.22	384.61	1.3449

h = kJ/kg, v = cm^3/g
and s = kJ/kg·K

THE PROPERTIES OF SATURATED DOWTHERM A

T	P	v_f	v_g	h_f	h_g
15.00				4.77	384.70
40.00				45.65	412.25
65.00				88.37	446.23
90.00		0.9836	60330.	133.16	481.33
105.00		1.0106	30457.	161.47	505.11
115.00	1.40	1.0161	18935.	180.62	519.42
125.00	1.80	1.0293	11837.	201.25	539.06
140.00	3.72	1.0369	7334.3	231.96	563.52
150.00	5.26	1.0514	4909.4	253.17	583.17
160.00	7.58	1.0613	3283.7	274.42	600.00
170.00	10.57	1.0680	2482.7	296.04	617.60
180.00	14.38	1.0732	1835.3	315.35	633.63
195.00	21.76	1.0932	1176.9	353.02	668.50
205.00	28.45	1.0987	910.04	378.14	689.78
215.00	36.04	1.1096	724.66	403.26	711.21
225.00	48.99	1.1256	544.14	428.28	727.36
240.00	66.05	1.1345	389.01	465.79	759.09
250.00	84.35	1.1417	311.37	491.48	780.56
260.00	103.38	1.1549	254.71	516.28	802.33
270.00	129.06	1.1750	209.57	540.76	820.25
280.00	154.82	1.1777	155.57	569.16	843.31
295.00	211.24	1.2058	119.63	609.53	874.94
305.00	251.31	1.2183	103.95	638.84	899.45
315.00	295.05	1.2356	92.530	668.14	924.62
325.00	338.49	1.2474	78.184	697.44	947.01
340.00	441.52	1.2720	59.728	741.40	985.10
350.00	519.32	1.2956	50.548	770.70	1007.0
360.00	606.48	1.3172	43.325	800.00	1030.2
370.00	699.05	1.3281	38.965	829.30	1055.5
380.00	802.10	1.3522	34.196	858.60	1077.3
400.00	1047.9	1.3986	26.686	917.21	1126.0

Datum - Sat. liquid h = 0 at 12°C.

Properties of Saturated Dichloromonofluoromethane (F-21)*

Temp., °F.	Abs. pressure, lb./sq. in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy, B.t.u./(lb.)(°R.)	
		Liquid	Vapor	Liquid	Vapor	Liquid	Vapor
t	p	v_f	v_g	h_f	h_g	s_f	s_g
−40	1.358	0.01058	32.09	0.00	114.56	0.0008	0.2730
−30	1.888	.01066	23.61	2.36	115.76	.0055	.2695
−20	2.578	.01075	17.66	4.71	116.96	.0109	.2663
−10	3.463	.01084	13.43	7.07	118.17	.0162	.2633
0	4.582	.01093	10.35	9.44	119.37	.0214	.2606
5	5.243	.01097	9.132	10.63	119.97	.0240	.2593
10	5.978	.01102	8.085	11.81	120.57	.0265	.2581
20	7.699	.01112	6.392	14.21	121.78	.0316	.2559
30	9.793	.01122	5.112	16.61	122.98	.0365	.2538
40	12.32	.01132	4.130	19.04	124.19	.0414	.2519
50	15.33	.01142	3.370	21.49	125.39	.0463	.2502
60	18.90	.01153	2.773	23.98	126.60	.0511	.2486
70	23.08	.01164	2.300	26.49	127.79	.0559	.2471
80	27.96	.01176	1.923	29.03	128.98	.0606	.2458
86	31.23	.01183	1.733	30.56	129.68	.0634	.2450
90	33.58	.01188	1.619	31.59	130.14	.0652	.2446
100	40.04	.01200	1.371	34.18	131.29	.0699	.2434
110	47.40	.01213	1.169	36.79	132.42	.0745	.2424
120	55.75	.01226	1.001	39.46	133.53	.0791	.2414
130	65.15	.01240	0.8623	42.13	134.61	.0837	.2405
140	75.72	.01254	.7457	44.86	135.66	.0882	.2396
150	87.51	.01269	.6476	47.62	136.68	.0927	.2388
160	100.6	.01284	.5646	50.43	137.69	.0972	.2381

* Courtesy Kinetic Chemicals, Inc.

Properties of Superheated Dichloromonofluoromethane (F-21)*

v, volume, cu. ft./lb.; h, enthalpy, B.t.u./lb.; s, entropy, B.t.u./(lb.)(°R.)
Parenthetic figures after pressures are saturation temperatures

Temp., °F.	Abs. pressure 1.2 lb./sq. in. (−43.6°F.)		
t	v	h	s
Sat.	36.02	114.13	0.2744
−40	36.34	114.57	.2754
−30	37.21	115.80	.2783
−20	38.08	117.03	.2812
−10	38.95	118.28	.2840
0	39.83	119.54	.2867
10	40.70	120.80	.2894
20	41.57	122.08	.2921
30	42.44	128.38	.2948
40	43.32	124.69	.2975
50	44.19	126.01	.3001
60	45.06	127.35	.3027
70	45.93	128.71	.3053
80	46.80	130.09	.3079
90	47.67	131.48	.3104
100	48.54	132.88	.3129
110	49.42	134.29	.3154
120	50.29	135.71	.3179
130	51.16	137.14	.3203
140	52.03	138.59	.3228
150	52.90	140.06	.3252
160	53.77	141.55	.3276
170	54.64	143.05	.3300
180	55.51	144.56	.3324
190	56.38	146.08	.3348
200	57.25	147.61	.3371
210	58.12	149.16	.3394
220	58.99	150.73	.3418
230	59.86	152.31	.3441
240	60.73	153.90	.3464
250	61.60	155.51	.3486

Temp., °F.	Abs. pressure 2 lb./sq. in. (−28.2°F.)		
t	v	h	s
Sat.	22.37	115.98	0.2689
−20	22.80	116.99	.2712
−10	23.33	118.24	.2740
0	23.85	119.49	.2768
10	24.38	120.76	.2795
20	24.90	122.04	.2822
30	25.43	123.34	.2849
40	25.95	124.65	.2876
50	26.47	125.98	.2902
60	27.00	127.33	.2928
70	27.52	128.69	.2954
80	28.05	130.07	.2979
90	28.57	131.45	.3005
100	29.09	132.85	.3030
110	29.62	134.26	.3055
120	30.14	135.68	.3080
130	30.66	137.12	.3104
140	31.19	138.57	.3129
150	31.71	140.03	.3153
160	32.23	141.52	.3177
170	32.76	143.02	.3201
180	33.28	144.53	.3225
190	33.80	146.05	.3249
200	34.33	147.59	.3272
210	34.85	149.14	.3296
220	35.37	150.71	.3319
230	35.89	152.29	.3342
240	36.42	153.88	.3365
250	36.94	155.59	.3388

Temp., °F.	Abs. pressure 4 lb./sq. in. (−4.9°F.)		
t	v	h	s
Sat.	11.74	118.78	0.2619
0	11.87	119.40	.2633
10	12.14	120.66	.2660
20	12.40	121.95	.2687
30	12.66	123.25	.2714
40	12.93	124.57	.2741
50	13.19	125.90	.2767
60	13.45	127.25	.2793
70	13.72	128.61	.2819
80	13.98	129.99	.2845
90	14.24	131.37	.2870
100	14.51	132.77	.2895
110	14.77	134.18	.2920
120	15.03	135.61	.2945
130	15.29	137.04	.2970
140	15.56	138.49	.2994
150	15.82	139.96	.3019
160	16.08	141.45	.3043
170	16.35	142.95	.3067
180	16.61	144.46	.3091
190	16.87	145.98	.3114
200	17.13	147.52	.3138
210	17.39	149.08	.3161
220	17.66	150.65	.3184
230	17.92	152.23	.3207
240	18.18	153.82	.3230
250	18.44	155.43	.3253
260	18.70	157.05	.3276
270	18.97	158.68	.3298

Temp., °F.	Abs. pressure 10 lb./sq. in. (30.9°F.)		
t	v	h	s
Sat.	5.014	123.10	0.2536
40	5.112	124.30	.2561
50	5.219	125.63	.2587
60	5.326	126.99	.2613
70	5.443	128.36	.2639
80	5.540	129.74	.2665
90	5.646	131.13	.2691
100	5.753	132.54	.2716
110	5.860	133.95	.2741
120	5.966	135.38	.2766
130	6.072	136.82	.2791
140	6.178	138.28	.2815
150	6.285	139.75	.2840
160	6.391	141.24	.2864
170	6.497	142.75	.2888
180	6.603	144.27	.2912
190	6.709	145.79	.2936
200	6.815	147.33	.2959
210	6.921	148.89	.2983
220	7.027	150.46	.3006
230	7.133	152.05	.3029
240	7.239	153.65	.3052
250	7.344	155.26	.3075
260	7.450	156.88	.3098
270	7.555	158.52	.3120
290	7.761	160.17	.3143
290	7.767	161.84	.3165
300	7.872	163.52	.3187
310	7.978	165.21	.3210

* Courtesy of Kinetic Chemicals, Inc.

Properties of Saturated Propane*

Temp., °F.	Abs. pressure, lb./sq.in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy, B.t.u./(lb.)(°R.)		Temp., °F.	Abs. pressure, lb./sq. in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy, B.t.u./(lb.)(°R.)	
		Liquid	Vapor	Liquid	Vapor	Liquid	Vapor			Liquid	Vapor	Liquid	Vapor	Liquid	Vapor
t	p	v_f	v_g	h_f	h_g	s_f	s_g	t	p	v_f	v_g	h_f	h_g	s_f	s_g
−80	5.65	0.0265	16.2	162.6	354.0	0.8794	1.3832	70	124.3	.03209	.854	245.7	394.4	1.0624	1.3427
−70	7.48	.0268	12.5	167.6	357.0	.8927	1.3781	80	143.6	.03269	.745	251.9	396.4	1.0737	1.3413
−60	9.78	.02703	9.77	172.7	360.0	.9060	1.3740	90	165.0	.03329	.643	258.2	398.3	1.0850	1.3400
−50	12.60	.02733	7.73	177.8	362.8	.9188	1.3702	100	188.7	.03390	.558	264.6	400.2	1.0963	1.3388
−40	16.00	.02763	6.16	183.0	365.7	.9315	1.3670	110	214.8	.03452	.487	271.1	401.9	1.1080	1.3378
−30	20.18	.02794	5.02	188.4	368.6	.9441	1.3640	120	243.4	.03532	.426	278.0	403.8	1.1195	1.3368
−20	25.05	.02826	4.06	193.8	371.5	.9568	1.3610	130	274.5	.03612	.370	285.2	405.4	1.1310	1.3356
−10	30.95	.02859	3.33	199.4	374.4	.9690	1.3582	140	308.4	.03702	.320	292.7	407.0	1.1430	1.3347
0	37.81	.02893	2.74	205.0	377.2	.9812	1.3555	150	345.4	.03817	.278	300.2	408.2	1.1552	1.3326
10	45.85	.02930	2.30	210.7	380.0	.9932	1.3531	160	385.0	.03962	.240	308.4	408.8	1.1680	1.3303
20	55.00	.02970	1.93	216.6	382.6	1.0050	1.3510	170	426.0	.04132	.208	317.5	408.6	1.1816	1.3272
30	65.70	.03011	1.60	222.3	385.1	1.0167	1.3491	180	473.2	.04367	.180	327.5	407.6	1.1970	1.3223
40	77.80	.03055	1.33	227.9	387.5	1.0283	1.3473	190	523.4	.04712	.149	339.2	404.6	1.2140	1.3156
50	91.50	.03101	1.14	233.8	389.9	1.0398	1.3456	200	575.0	.0521	.113	353.5	398.3	1.2360	1.3040
60	106.9	.03150	0.984	239.6	392.2	1.0511	1.3441								

* Stearns and George, *Ind. Eng. Chem.*, **35**, 602 (1943); with permission.

Properties of Superheated Propane*

v, volume, cu. ft./lb.; h, enthalpy, B.t.u./lb.; s, entropy, B.t.u./(lb.)(°R.); p, absolute pressure, lb./sq. in.
Parenthetic figures after pressures are saturation temperatures

Temp., °F.	7.35 (−70.7°F.)			12.24 (−51°F.)			14.696 (−43.708°F.)			20 (−30.30°F.)		
	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	12.72	356.5	1.3785	7.90	362.75	1.3718	6.66	364.6	1.3681	5.050	368.4	1.3640
−60	13.11	360.3	1.3869									
−40	13.79	366.8	1.4023	8.182	366.2	1.3796	6.775	365.8	1.3710	......	......	......
−20	14.47	373.4	1.4177	8.591	373.0	1.3948	7.123	372.8	1.3866	5.186	372.2	1.3719
0	15.14	380.2	1.4329	9.000	379.9	1.4100	7.471	379.8	1.4020	5.439	379.3	1.3873
20	15.82	387.4	1.4481	9.409	387.0	1.4252	7.816	386.8	1.4772	5.695	386.5	1.4025
40	16.50	394.8	1.4633	9.818	394.4	1.4404	8.160	394.3	1.4324	5.951	393.9	1.4177
60	17.17	402.4	1.4785	10.23	402.0	1.4556	8.503	401.9	1.4474	6.207	401.6	1.4327
80	17.85	410.2	1.4937	10.64	410.0	1.4706	8.844	409.8	1.4623	6.461	409.4	1.4477
100	18.52	418.4	1.5088	11.050	418.1	1.4854	9.187	418.0	1.4772	6.716	417.6	1.4625
120	19.20	426.6	1.5235	11.45	426.2	1.5002	9.528	426.2	1.4918	6.969	425.9	1.4771
140	19.88	435.2	1.5380	11.86	435.9	1.5148	9.869	434.8	1.5064	7.221	434.6	1.4917
160	20.55	444.1	1.5524	12.27	443.8	1.5293	10.21	443.7	1.5210	7.473	443.5	1.5063
180	21.23	453.1	1.5668	12.68	452.9	1.5437	10.51	452.9	1.5256	7.725	452.6	1.5209
200	21.90	462.5	1.5812	13.09	462.4	1.5581	10.84	462.4	1.5500	7.977	461.9	1.5355

Temp., °F.	30 (−11.52°F.)			40 (+2.90°F.)			60 (24.80°F.)			80 (41.69°F.)		
	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	3.30	374.00	1.3588	2.61	378.2	1.3548	1.76	383.8	1.3501	1.32	387.8	1.3470
0	3.559	378.3	1.3678	......	......	......	......	......	......	......	......	......
20	3.735	385.7	1.3830	2.753	384.7	1.3684	......	......	......	......	......	......
40	3.911	393.1	1.3980	2.889	392.2	1.3838	1.863	390.1	1.3630	......	......	......
60	4.087	400.8	1.4130	3.025	400.0	1.3990	1.959	398.2	1.3789	1.424	396.1	1.3624
80	4.261	408.7	1.4280	3.159	408.0	1.4140	2.053	406.3	1.3948	1.500	404.6	1.3785
100	4.432	417.0	1.4428	3.289	416.2	1.4290	2.145	414.8	1.4102	1.573	413.2	1.3940
120	4.602	425.4	1.4576	3.419	424.6	1.4440	2.235	423.4	1.4254	1.644	421.8	1.4094
140	4.772	434.0	1.4724	3.549	433.4	1.4588	2.325	432.2	1.4404	1.714	430.9	1.4248
160	4.942	443.0	1.4872	3.679	442.5	1.4736	2.415	441.4	1.4554	1.784	440.3	1.4402
180	5.112	452.2	1.5020	3.809	451.8	1.4884	2.505	450.8	1.4703	1.852	449.8	1.4554
200	5.282	461.8	1.5168	3.939	461.3	1.5030	2.593	460.5	1.4851	1.920	459.7	1.4704

Temp., °F.	100 (55.62°F.)			130 (73.20°F.)			160 (87.71°F.)		
	v	h	s	v	h	s	v	h	s
Sat.	1.06	391.2	1.3448	0.8165	395.0	1.3424	0.6685	397.9	1.3404
60	1.094	395.5	1.3488	......	......	......	......	......	......
80	1.164	402.6	1.3656	.8456	398.8	1.3486	......	......	......
100	1.227	411.4	1.3816	.9045	408.3	1.3659	.6969	404.7	1.3521
120	1.289	420.3	1.3962	.9588	417.8	1.3822	.7464	414.6	1.3698
140	1.347	429.6	1.4130	1.006	427.9	1.3987	.7908	424.8	1.3867
160	1.400	439.1	1.4286	1.052	437.2	1.4150	.8319	434.9	1.4031
180	1.460	448.8	1.4440	1.098	447.0	1.4310	.8712	445.1	1.4195
200	1.516	458.8	1.4598	1.143	457.1	1.4468	.9093	455.4	1.4357

Temp., °F.	190 (100.50°F.)			220 (111.85°F.)			250 (122.12°F.)			300 (137.55°F.)		
	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	0.5540	400.4	1.3388	0.4738	402.5	1.3375	0.4130	404.2	1.3355	0.3332	406.6	1.3345
120	.5995	411.6	1.3580	.4911	407.7	1.3460	......	......	......	......	......	......
140	.6415	422.1	1.3759	.5314	419.2	1.3650	.4473	415.9	1.3550	.3392	408.7	1.3372
160	.6792	432.4	1.3930	.5673	429.8	1.3827	.4816	427.1	1.3732	.3745	422.0	1.3580
180	.7144	443.0	1.4096	.5998	440.6	1.3999	.5121	438.0	1.3908	.4037	433.2	1.3765
200	.7472	453.4	1.4255	.6302	451.2	1.4161	.5408	448.9	1.4074	.4303	444.9	1.3944

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Properties of Saturated Dowtherm A*

(73.5 % diphenyloxide, 26.5 % diphenyl)

Temp., °F. t	Abs. pressure, lb./sq. in. p	Volume, cu. ft./lb. Liquid v_f	Volume, cu. ft./lb. Vapor v_g	Enthalpy, B.t.u./lb. Liquid h_f	Enthalpy, B.t.u./lb. Vapor h_g
53.6	......	......	......	0	164
100	......	......	......	18.0	176
150	......	......	......	38.4	192
200	......	0.0160	833	60.0	210
220	......	.0162	500	69.0	217
240	0.20	.0163	294	78.2	224
260	.29	.0165	179	87.7	232
280	.49	.0166	125	97.5	240
300	.74	.0168	83.3	108.0	250
320	1.1	.0170	52.6	118.0	258
340	1.6	.0171	38.5	128.0	266
360	2.2	.0173	27.8	138.0	275
380	3.0	.0175	20.0	150.0	286
400	4.1	.0176	14.7	162.0	296
420	5.4	.0178	11.4	174.0	306
440	6.9	.0180	8.40	186.0	316
460	8.8	.0181	6.54	198.0	326
480	12	.0183	5.13	210.0	334
500	15	.0185	4.08	222.0	345
520	19	.0188	3.23	234.0	354
540	24	.0190	2.44	247.0	365
560	30	.0193	2.00	260.0	375
580	36	.0195	1.67	274.0	386
600	43	.0198	1.47	288.0	398
620	51	.0201	1.20	302.0	409
640	62	.0204	0.980	316.0	421
660	74	.0207	.826	330.0	432
680	88	.0211	.694	344.0	443
700	103	.0213	.617	358.0	455
720	120	.0218	.526	372.0	465
750	150	.0225	.417	393.0	482

* Dow Chemical Co.

Properties of Saturated Ammonia*

Temp., °F. t	Abs. pressure, lb./sq. in. p	Volume, cu. ft./lb. Liquid v_f	Volume, cu. ft./lb. Vapor v_g	Enthalpy, B.t.u./lb. Liquid h_f	Enthalpy, B.t.u./lb. Vapor h_g	Entropy, B.t.u./(lb.)(°R.) Liquid s_f	Entropy, B.t.u./(lb.)(°R.) Vapor s_g
−60	5.55	0.02278	44.73	−21.2	589.6	−0.0517	1.4769
−50	7.67	.02299	33.08	−10.6	593.7	− .0256	1.4497
−40	10.41	.02322	24.86	0.0	597.6	.0000	1.4242
−30	13.90	....	18.97	10.7	601.4	.0250	1.4001
−20	18.30	.02369	14.68	21.4	605.0	.0497	1.3774
−16	20.34	....	13.29	25.6	606.4	.0594	1.3686
−12	22.56	....	12.06	30.0	607.8	.0690	1.3600
− 8	24.97	....	10.97	34.3	609.2	.0786	1.3516
− 4	27.59	....	9.991	38.6	610.5	.0880	1.3433
0	30.42	.02419	9.116	42.9	611.8	.0975	1.3352
4	33.47	....	8.333	47.2	613.0	.1069	1.3273
8	36.77	....	7.629	51.6	614.3	.1162	1.3195
12	40.31	....	6.996	56.0	615.5	.1254	1.3118
16	44.12	....	6.425	60.3	616.6	.1346	1.3043
20	48.21	.02474	5.910	64.7	617.8	.1437	1.2969
24	52.59	....	5.443	69.1	618.9	.1528	1.2897
28	57.28	....	5.021	73.5	619.9	.1618	1.2825
32	62.29	....	4.637	77.9	621.0	.1708	1.2755
36	67.63	....	4.289	82.3	622.0	.1797	1.2686
40	73.32	.02533	3.971	86.8	623.0	.1885	1.2618
50	89.19	.02564	3.294	97.9	625.2	.2105	1.2453
60	107.6	.02597	2.751	109.2	627.3	.2322	1.2294
70	128.8	.02632	2.312	120.5	629.1	.2537	1.2140
80	153.0	.02668	1.955	132.0	630.7	.2749	1.1991
90	180.6	.02707	1.661	143.5	632.0	.2958	1.1846
100	211.9	.02747	1.419	155.2	633.0	.3166	1.1705
110	247.0	.02790	1.217	167.0	633.7	.3372	1.1566
120	286.4	.02836	1.047	179.0	634.0	.3576	1.1427
125	307.8	.02860	0.973	183.9	634.0	.3659	1.1372

* *U. S. Bur. Standards Circ.* 142, 1923.

Table 204. Properties of Superheated Ammonia*

v, volume, cu. ft./lb.; h, enthalpy, B.t.u /lb.; s, entropy, B.t.u./(lb.)(°R.)
Absolute pressure, lb. per sq. in. (saturation temperature, °F., in parentheses)

Temp., °F.	5 (−63.11)			7 (−52.88°)			10 (−41.34)			14 (−29.76)			18 (−20.61)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	49.31	588.5	1.4857	36.01	592.5	1.4574	25.81	597.1	1.4276	18.85	601.4	1.3996	14.90	604.8	1.3787
−50	51.05	595.2	1.5025	36.29	594.0	1.4611									
−40	52.36	600.3	1.5149	37.25	599.3	1.4739									
−30	53.67	605.4	1.5269	38.19	604.5	1.4861	26.58	603.2	1.4420						
−20	54.97	610.4	1.5385	39.13	609.6	1.4979	27.26	608.5	1.4542	19.33	606.8	1.4119	14.93	605.1	1.3795
−10	56.26	615.4	1.5498	40.07	614.7	1.5094	27.92	613.7	1.4659	19.82	612.2	1.4241	15.32	610.7	1.3921
0	57.55	620.4	1.5608	41.00	619.8	1.5206	28.58	618.9	1.4773	20.30	617.6	1.4358	15.70	616.2	1.4042
10	58.84	625.4	1.5716	41.93	624.9	1.5314	29.24	624.0	1.4884	20.78	622.8	1.4472	16.08	621.6	1.4158
20	60.12	630.4	1.5821	42.85	629.9	1.5421	29.90	629.1	1.4992	21.26	628.0	1.4582	16.46	626.9	1.4270
30	61.41	635.4	1.5925	43.77	635.0	1.5525	30.55	634.2	1.5097	21.73	633.2	1.4688	16.83	632.2	1.4380
40	62.69	640.4	1.6026	44.69	640.0	1.5627	31.20	639.3	1.5200	22.20	638.4	1.4793	17.20	637.5	1.4486
50	63.96	645.5	1.6125	45.61	645.0	1.5727	31.85	644.4	1.5301	22.67	643.6	1.4896	17.57	642.7	1.4590
60	65.24	650.5	1.6223	46.53	650.1	1.5825	32.49	649.5	1.5400	23.14	648.7	1.4996	17.94	647.9	1.4691
70	66.51	655.5	1.6319	47.44	655.2	1.5921	33.14	654.6	1.5497	23.60	653.9	1.5094	18.30	653.1	1.4790
80	67.79	660.6	1.6413	48.36	660.2	1.6016	33.78	659.7	1.5593	24.06	659.0	1.5191	18.67	658.4	1.4887

* *U. S. Bur. Standards Circ.* 142, 1923.

Properties of Superheated Ammonia*—(*Continued*)

Temp. °F.	5 (−63.11)			7 (−52.88°)			10 (−41.34)			14 (−29.76)			18 (−20.61)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
90	69.06	665.6	1.6506	49.27	665.3	1.6110	34.42	664.8	1.5687	24.53	664.2	1.5285	19.03	663.6	1.4983
100	70.33	670.7	1.6598	50.18	670.4	1.6202	35.07	670.0	1.5779	24.99	669.4	1.5378	19.39	668.8	1.5077
110	71.60	675.8	1.6689	51.09	675.5	1.6292	35.71	675.1	1.5870	25.45	674.5	1.5470	19.75	674.0	1.5169
120	72.87	680.9	1.6778	52.00	680.7	1.6382	36.35	680.3	1.5960	25.91	679.7	1.5560	20.11	679.2	1.5260
130	74.14	686.1	1.6865	52.91	685.8	1.6470	36.99	685.4	1.6049	26.37	684.9	1.5649	20.47	684.4	1.5349
140	75.41	691.2	1.6952	53.82	691.0	1.6557	37.62	690.6	1.6136	26.83	690.1	1.5737	20.83	689.7	1.5438
150	76.68	696.4	1.7038	54.73	696.2	1.6643	38.26	695.8	1.6222	27.29	695.4	1.5824	21.19	694.9	1.5525
160	77.95	701.6	1.7122	55.63	701.4	1.6727	38.90	701.1	1.6307	27.74	700.6	1.5909	21.54	700.2	1.5610
170	79.21	706.8	1.7206	56.54	706.6	1.6811	39.54	706.3	1.6391	28.20	705.9	1.5993	21.90	705.5	1.5695
180	80.48	712.1	1.7289	57.45	711.9	1.6894	40.17	711.6	1.6474	28.66	711.2	1.6076	22.26	710.8	1.5778
190	.....	.....	.....	.....	.....	.....	40.81	716.9	1.6556	29.11	716.5	1.6159	22.61	716.1	1.5861
200	.....	.....	.....	.....	.....	.....	41.45	722.2	1.6637	29.57	721.8	1.6240	22.97	721.4	1.5943
220	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	23.68	732.2	1.6103

Temp., °F.	24 (−9.58)			30 (−0.57)			38 (9.42)			48 (19.80)		
	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	11.39	608.6	1.3549	9.236	611.6	1.3364	7.396	614.7	1.3168	5.934	617.7	1.2973
0	11.67	614.1	1.3670	9.250	611.9	1.3371						
10	11.96	619.7	1.3791	9.492	617.8	1.3497	7.407	615.0	1.3175			
20	12.25	625.2	1.3907	9.731	623.5	1.3618	7.603	621.0	1.3301	5.937	617.8	1.2976
30	12.54	630.7	1.4019	9.966	629.1	1.3733	7.795	626.9	1.3422	6.096	624.0	1.3103
40	12.82	636.1	1.4128	10.20	634.6	1.3845	7.983	632.6	1.3538	6.251	630.0	1.3225
50	13.11	641.4	1.4234	10.43	640.1	1.3953	8.170	638.3	1.3650	6.404	635.9	1.3341
60	13.39	646.7	1.4337	10.65	645.5	1.4059	8.353	643.8	1.3758	6.554	641.6	1.3453
70	13.66	652.0	1.4438	10.88	650.9	1.4161	8.535	649.3	1.3863	6.702	647.3	1.3561
80	13.94	657.3	1.4537	11.10	656.2	1.4261	8.716	654.8	1.3965	6.848	652.9	1.3666
90	14.22	662.6	1.4634	11.33	661.6	1.4359	8.895	660.2	1.4065	6.993	658.5	1.3768
100	14.49	667.8	1.4729	11.55	666.9	1.4456	9.073	665.6	1.4163	7.137	664.0	1.3868
110	14.76	673.1	1.4822	11.77	672.2	1.4550	9.250	671.0	1.4258	7.280	669.5	1.3965
120	15.04	678.4	1.4914	11.99	677.5	1.4642	9.426	676.4	1.4352	7.421	675.0	1.4061
130	15.31	683.6	1.5004	12.21	682.9	1.4733	9.602	681.8	1.4444	7.562	680.5	1.4154
140	15.58	688.9	1.5093	12.43	688.2	1.4823	9.776	687.2	1.4534	7.702	685.9	1.4246
150	15.85	694.2	1.5180	12.65	693.5	1.4911	9.950	692.6	1.4623	7.842	691.4	1.4336
160	16.12	699.5	1.5266	12.87	698.8	1.4998	10.12	698.0	1.4711	7.981	696.8	1.4425
170	16.39	704.8	1.5352	13.08	704.2	1.5083	10.30	703.3	1.4797	8.119	702.3	1.4512
180	16.66	710.2	1.5436	13.30	709.6	1.5168	10.47	708.7	1.4883	8.257	707.7	1.4598
190	16.93	715.5	1.5518	13.52	714.9	1.5251	10.64	714.2	1.4966	8.395	713.2	1.4683
200	17.20	720.9	1.5600	13.73	720.3	1.5334	10.81	719.6	1.5049	8.532	718.7	1.4766
220	17.73	731.7	1.5761	14.16	731.1	1.5495	11.16	730.5	1.5212	8.805	729.6	1.4930
240	18.27	742.6	1.5919	14.59	742.0	1.5653	11.50	741.4	1.5371	9.077	740.6	1.5090
260	.....	.....	.....	15.02	753.0	1.5808	11.84	752.4	1.5526	9.348	751.7	1.5246
280	.....	.....	.....	.....	.....	.....	12.18	763.5	1.5678	9.619	762.9	1.5399
300	.....	.....	.....	.....	.....	.....	.....	.....	.....	9.888	774.1	1.5548

Temp., °F.	60 (30.21)			80 (44.40)			100 (56.05)			120 (66.02)		
	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	4.805	620.5	1.2787	3.655	624.0	1.2545	2.952	626.5	1.2356	2.476	628.4	1.2201
30												
40	4.933	626.8	1.2913	.....	.....	.....	.....	.....	.....	.....	.....	.....
50	5.060	632.9	1.3035	3.712	627.7	1.2619	.....	.....	.....	.....	.....	.....
60	5.184	639.0	1.3152	3.812	634.3	1.2745	2.985	629.3	1.2409	.....	.....	.....
70	5.307	644.9	1.3265	3.909	640.6	1.2866	3.068	636.0	1.2539	2.505	631.3	1.2255
80	5.428	650.7	1.3373	4.005	646.7	1.2981	3.149	642.6	1.2661	2.576	638.3	1.2386
90	5.547	656.4	1.3479	4.098	652.8	1.3092	3.227	649.0	1.2778	2.645	645.0	1.2510
100	5.665	662.1	1.3581	4.190	658.7	1.3199	3.304	655.2	1.2891	2.712	651.6	1.2628
110	5.781	667.7	1.3681	4.281	664.6	1.3303	3.380	661.3	1.2999	2.778	658.0	1.2741
120	5.897	673.3	1.3778	4.371	670.4	1.3404	3.454	667.3	1.3104	2.842	664.2	1.2850
130	6.012	678.9	1.3873	4.460	676.1	1.3502	3.527	673.3	1.3206	2.905	670.4	1.2956
140	6.126	684.4	1.3966	4.548	681.8	1.3598	3.600	679.2	1.3305	2.967	676.5	1.3058
150	6.239	689.9	1.4058	4.635	687.5	1.3692	3.672	685.0	1.3401	3.029	682.5	1.3157
160	6.352	695.5	1.4148	4.722	693.2	1.3784	3.743	690.8	1.3495	3.089	688.4	1.3254
170	6.464	701.0	1.4236	4.808	698.8	1.3874	3.813	696.6	1.3588	3.149	694.3	1.3348
180	6.576	706.5	1.4323	4.893	704.4	1.3963	3.883	702.3	1.3678	3.209	700.2	1.3441
190	6.687	712.0	1.4409	4.978	710.0	1.4050	3.952	708.0	1.3767	3.268	706.0	1.3531
200	6.798	717.5	1.4493	5.063	715.6	1.4136	4.021	713.7	1.3854	3.326	711.8	1.3620
210	6.909	723.1	1.4576	5.147	721.3	1.4220	4.090	719.4	1.3940	3.385	717.6	1.3707
220	7.019	728.6	1.4658	5.231	726.9	1.4304	4.158	725.1	1.4024	3.442	723.1	1.3793
230	.....	.....	.....	5.315	732.5	1.4386	4.226	730.8	1.4108	3.500	729.2	1.3877
240	7.238	739.7	1.4819	5.398	738.1	1.4467	4.294	736.5	1.4190	3.557	734.9	1.3960
250	.....	.....	.....	5.482	743.8	1.4547	4.361	742.2	1.4271	3.614	740.7	1.4042
260	7.457	750.9	1.4976	5.565	749.4	1.4626	4.428	747.9	1.4350	3.671	746.5	1.4123
270	.....	.....	.....	.....	.....	.....	.....	.....	.....	3.727	752.2	1.4202
280	7.675	762.1	1.5130	5.730	760.7	1.4781	4.562	759.4	1.4507	3.783	758.0	1.4281
290	.....	.....	.....	.....	.....	.....	.....	.....	.....	3.839	763.8	1.4359
300	7.892	773.3	1.5281	5.894	772.1	1.4933	4.695	770.8	1.4660	3.895	769.6	1.4435

Properties of Superheated Ammonia*—(*Concluded*)

Temp., °F.	140 (74.79)			160 (82.64)			180 (89.78)		
	v	h	s	v	h	s	v	h	s
Sat.	2.132	629.9	1.2068	1.872	631.1	1.1952	1.667	632.0	1.1850
80	2.166	633.8	1.2140						
90	2.228	640.9	1.2272	1.914	636.6	1.2055	1.668	632.2	1.1853
100	2.288	647.8	1.2396	1.969	643.9	1.2186	1.720	639.9	1.1992
110	2.347	654.5	1.2515	2.023	651.0	1.2311	1.770	647.3	1.2123
120	2.404	661.1	1.2628	2.075	657.8	1.2429	1.818	654.4	1.2247
130	2.460	667.4	1.2738	2.125	664.4	1.2542	1.865	661.3	1.2364
140	2.515	673.7	1.2843	2.175	670.9	1.2652	1.910	668.0	1.2477
150	2.569	679.9	1.2945	2.224	677.2	1.2575	1.955	674.6	1.2586
160	2.622	686.0	1.3045	2.272	683.5	1.2859	1.999	681.0	1.2691
170	2.675	692.0	1.3141	2.319	689.7	1.2958	2.042	687.3	1.2792
180	2.727	698.0	1.3236	2.365	695.8	1.3054	2.084	693.6	1.2891
190	2.779	704.0	1.3328	2.411	701.9	1.3148	2.126	699.8	1.2987
200	2.830	709.9	1.3418	2.457	707.9	1.3240	2.167	705.9	1.3081
210	2.880	715.8	1.3507	2.502	713.9	1.3331	2.208	712.0	1.3172
220	2.931	721.6	1.3594	2.547	719.9	1.3419	2.248	718.1	1.3262
230	2.981	727.5	1.3679	2.591	725.8	1.3506	2.288	724.1	1.3350
240	3.030	733.3	1.3763	2.635	731.7	1.3591	2.328	730.1	1.3436
250	3.080	739.2	1.3846	2.679	737.6	1.3675	2.367	736.1	1.3521
260	3.129	745.0	1.3928	2.723	743.5	1.3757	2.407	732.0	1.3605
270	3.179	750.8	1.4008	2.766	749.4	1.3838	2.446	748.0	1.3687
280	3.227	756.7	1.4088	2.809	755.3	1.3919	2.484	753.9	1.3768
290	3.275	762.5	1.4166	2.852	761.2	1.3998	2.523	759.9	1.3847
300	3.323	768.3	1.4243	2.895	767.1	1.4076	2.561	765.8	1.3926
320	3.420	780.0	1.4395	2.980	778.9	1.4229	2.637	777.7	1.4081
340	.....	.....	......	3.064	790.7	1.4379	2.713	789.6	1.4231

Temp., °F.	200 (96.34)			220 (102.42)			240 (108.09)			260 (113.42)			300 (123.21)		
	v	h	s	v	h	s	v	h	s	v	h	s	v	h	s
Sat.	1.502	632.7	1.1756	1.367	633.2	1.1671	1.253	633.6	1.1592	1.155	633.9	1.1518	0.999	634.0	1.1383
110	1.567	643.4	1.1947	1.400	639.4	1.1781	1.261	635.3	1.1621						
120	1.612	650.9	1.2077	1.443	647.3	1.1917	1.302	643.5	1.1764	1.182	639.5	1.1617			
130	1.656	658.1	1.2200	1.485	654.8	1.2045	1.342	651.3	1.1898	1.220	647.8	1.1757	1.023	640.1	1.1487
140	1.698	665.0	1.2317	1.525	662.0	1.2167	1.380	658.8	1.2025	1.257	655.6	1.1889	1.058	648.7	1.1632
150	1.740	671.8	1.2429	1.564	669.0	1.2281	1.416	666.1	1.2145	1.292	663.1	1.2014	1.091	656.9	1.1767
160	1.780	678.4	1.2537	1.601	675.8	1.2394	1.452	673.1	1.2259	1.326	670.4	1.2132	1.123	664.7	1.1894
170	1.820	684.9	1.2641	1.638	682.5	1.2501	1.487	680.0	1.2369	1.359	677.5	1.2245	1.153	672.2	1.2014
180	1.859	691.3	1.2742	1.675	689.1	1.2604	1.521	686.7	1.2475	1.391	684.4	1.2354	1.183	679.5	1.2129
190	1.897	697.7	1.2840	1.710	695.5	1.2704	1.554	693.3	1.2577	1.422	691.1	1.2458	1.211	686.5	1.2239
200	1.935	703.9	1.2935	1.745	701.9	1.2801	1.587	699.8	1.2677	1.453	697.7	1.2560	1.239	693.5	1.2344
210	1.972	710.1	1.3029	1.780	708.2	1.2896	1.619	706.2	1.2773	1.484	704.3	1.2658	1.267	700.3	1.2447
220	2.009	716.3	1.3120	1.814	714.4	1.2989	1.651	712.6	1.2867	1.514	710.7	1.2754	1.294	706.9	1.2546
230	2.046	722.4	1.3209	1.848	720.6	1.3079	1.683	718.9	1.2959	1.543	717.1	1.2847	1.320	713.5	1.2642
240	2.082	728.4	1.3296	1.881	726.8	1.3168	1.714	725.1	1.3049	1.572	723.4	1.2938	1.346	720.0	1.2736
250	2.118	734.5	1.3382	1.914	732.9	1.3255	1.745	731.3	1.3137	1.601	729.7	1.3027	1.372	726.5	1.2827
260	2.154	740.5	1.3467	1.947	739.0	1.3340	1.775	737.5	1.3224	1.630	736.0	1.3115	1.397	732.9	1.2917
270	2.189	746.5	1.3550	1.980	745.1	1.3424	1.805	743.6	1.3308	1.658	742.2	1.3200	1.422	739.2	1.3004
280	2.225	752.5	1.3631	2.012	751.1	1.3507	1.835	749.8	1.3392	1.686	748.4	1.3285	1.447	745.5	1.3090
290	2.260	758.5	1.3712	2.044	757.2	1.3588	1.865	755.9	1.3474	1.714	754.5	1.3367	1.472	751.8	1.3175
300	2.295	764.5	1.3791	2.076	763.2	1.3668	1.895	762.0	1.3554	1.741	760.7	1.3449	1.496	758.1	1.3257
320	2.364	776.5	1.3947	2.140	775.3	1.3825	1.954	774.1	1.3712	1.796	772.9	1.3608	1.544	770.5	1.3419
340	2.432	788.5	1.4099	2.203	787.4	1.3978	2.012	786.3	1.3866	1.850	785.2	1.3763	1.592	782.9	1.3576
360	2.500	800.5	1.4247	2.265	799.5	1.4127	2.069	798.4	1.4016	1.904	797.4	1.3914	1.639	795.3	1.3729
380	2.568	812.5	1.4392	2.327	811.6	1.4273	2.126	810.6	1.4163	1.957	809.6	1.4062	1.686	807.7	1.3878

Properties of Saturated Ethylene*

Temp., °F.	Abs. pressure, atm.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy, B.t.u./(lb.)(°R.)	
		Liquid	Vapor	Liquid	Vapor	Liquid	Vapor
t	p	v_f	v_g	h_f	h_g	s_f	s_g
−272.47	0.0012	....	4064.0	−68.19	176.5	−0.2826	1.024
−260.0	.0037	....	1405.0	−60.84	180.1	−.2351	0.972
−240.0	.0177	....	328.6	−49.16	185.7	−.1887	.880
−220.0	.0606	....	103.1	−37.56	191.4	−.1382	.817
−200.0	.169	....	39.74	−26.03	197.1	−.0922	.767
−180.0	.402	....	17.92	−14.57	202.6	−.0498	.727
−160.0	.837	....	9.047	−3.12	207.2	−.0103	.692
−154.66	1.0000	0.02818	7.6712	0.0	207.9	.0000	.6814
−140.00	1.5775	.02877	5.005	+8.6	210.5	+.0249	.6563
−120.00	2.7376	.02964	2.987	20.8	214.2	.0648	.6340
−100.00	4.4616	.03122	1.879	32.8	217.2	.0995	.6121
−80.00	6.8697	.03179	1.732	45.2	219.9	.1374	.5974
−60.00	10.099	.03308	0.857	57.9	221.9	.1666	.5769
−40.00	14.0338	.03468	.593	70.8	222.8	.1935	.5556
−20.00	19.722	.03662	.419	84.7	222.7	.2245	.5383
0.00	26.397	.03912	.301	100.3	221.0	.2577	.5202
+20.00	34.55	.04292	.212	119.6	216.6	.2968	.4991
+40.00	44.54	.05035	.139	148.7	206.2	.3533	.4683
+49.82	50.50	.070	.070	171.8	171.8	.3964	.3964

* York and White, *Trans. Am. Inst. Chem. Engrs.*, **40**, 227 (1944).

Properties of Superheated Ethylene*

v, volume, cu. ft./lb.; h, enthalpy, B.t.u./lb.; s, entropy, B.t.u./(lb.)(°R.)

Abs. pressure, atm.	Temperature, °F.																			
	−140°	−120°	−100°	−80°	−60°	−40°	−20°	0°	20°	40°	60°	100°	140°	180°	220°	260°	320°	380°	440°	500°
v	8.061	8.617	9.168	9.714	10.256	10.794	11.329	11.862	12.497	12.924	13.453	14.51	15.56	16.61	17.66	18.71	20.28	21.85	23.42	24.99
1 h	212.2	218.3	224.3	230.6	236.9	243.3	249.9	256.6	263.4	270.4	277.7	292.6	308.3	324.8	342.0	360.0	388.4	418.4	450.4	482.9
s	0.695	0.714	0.732	0.748	0.764	0.780	0.795	0.810	0.825	0.839	0.853	0.881	0.908	0.934	0.961	0.986	1.025	1.062	1.098	1.133
v	.....	4.190	4.479	4.765	5.047	5.324	5.599	5.873	6.144	6.414	6.683	7.217	7.749	8.280	8.808	9.34	10.12	10.91	11.70	12.48
2 h	.....	215.9	222.3	228.9	235.4	242.0	248.7	255.6	262.5	269.5	276.9	291.9	307.6	324.2	341.5	359.7	388.1	418.2	450.2	482.7
s	.....	0.659	0.677	0.693	0.710	0.726	0.741	0.756	0.771	0.785	0.799	0.828	0.855	0.881	0.909	0.934	0.974	1.011	1.048	1.083
v	.....	.....	2.130	2.287	2.439	2.588	2.733	2.878	3.019	3.159	3.297	3.571	3.843	4.113	4.380	4.647	5.046	5.443	5.838	6.234
4 h	.....	.....	218.2	225.5	232.4	239.3	246.3	253.4	260.6	267.8	275.3	290.6	306.6	323.3	340.6	358.9	387.4	417.6	449.7	482.3
s	.....	.....	0.623	0.641	0.658	0.675	0.691	0.706	0.722	0.737	0.751	0.780	0.807	0.833	0.860	0.886	0.925	0.963	1.000	1.035
v	.....	.....	.....	1.456	1.566	1.673	1.776	1.833	1.877	2.099	2.168	2.356	2.541	2.724	2.905	3.084	3.352	3.620	3.885	4.150
6 h	.....	.....	.....	221.6	229.1	236.6	243.9	251.2	258.6	266.0	273.7	289.3	305.4	322.3	339.8	358.1	386.8	417.0	449.2	481.8
s	.....	.....	.....	0.607	0.625	0.642	0.659	0.675	0.691	0.706	0.721	0.750	0.777	0.804	0.831	0.856	0.896	0.934	0.971	1.006
v	.....	.....	.....	.....	0.860	0.936	1.006	1.074	1.140	1.202	1.264	1.382	1.499	1.613	1.724	1.834	1.999	2.161	2.323	2.483
10 h	.....	.....	.....	.....	222.4	230.8	238.8	246.6	254.4	262.4	270.4	286.5	303.1	320.3	338.0	356.5	385.4	415.8	448.2	480.9
s	.....	.....	.....	.....	0.578	0.599	0.616	0.632	0.649	0.665	0.680	0.711	0.739	0.766	0.794	0.820	0.860	0.898	0.934	0.969
v	.....	.....	.....	.....	.....	.....	0.614	0.669	0.718	0.765	0.811	0.897	0.978	1.057	1.134	1.209	1.322	1.433	1.542	1.650
15 h	.....	.....	.....	.....	.....	.....	230.9	239.9	248.7	257.5	266.2	283.0	300.1	317.7	335.8	354.6	383.8	414.4	447.0	479.9
s	.....	.....	.....	.....	.....	.....	0.573	0.593	0.612	0.630	0.646	0.677	0.707	0.734	0.763	0.789	0.829	0.867	0.903	0.938
v	.....	.....	.....	.....	.....	.....	.....	0.459	0.503	0.545	0.583	0.652	0.717	0.778	0.838	0.897	0.983	1.068	1.151	1.233
20 h	.....	.....	.....	.....	.....	.....	.....	232.5	242.5	252.1	261.3	279.2	296.9	315.1	333.5	352.6	382.1	413.0	445.7	478.8
s	.....	.....	.....	.....	.....	.....	.....	0.560	0.582	0.591	0.618	0.651	0.682	0.710	0.740	0.766	0.807	0.845	0.882	0.917
v	.....	.....	.....	.....	.....	.....	.....	.....	0.282	0.318	0.350	0.405	0.454	0.498	0.543	0.585	0.645	0.704	0.760	0.817
30 h	.....	.....	.....	.....	.....	.....	.....	.....	226.9	239.5	250.8	271.2	290.2	309.5	328.7	348.3	378.6	410.1	443.3	476.7
s	.....	.....	.....	.....	.....	.....	.....	.....	0.528	0.552	0.574	0.613	0.646	0.676	0.706	0.734	0.775	0.814	0.851	0.887
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.192	0.228	0.279	0.323	0.359	0.395	0.428	0.476	0.521	0.565	0.609
40 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	221.1	237.3	262.3	283.4	303.7	323.8	344.1	375.1	407.2	440.8	474.5
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.503	0.534	0.580	0.617	0.650	0.681	0.709	0.751	0.791	0.828	0.865
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.138	0.201	0.243	0.275	0.309	0.334	0.375	0.412	0.448	0.484
50 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	218.1	250.3	275.6	297.5	318.7	339.7	371.5	404.2	438.2	472.4
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.484	0.550	0.590	0.626	0.659	0.688	0.731	0.772	0.810	0.847
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.069	0.144	0.186	0.219	0.248	0.273	0.307	0.340	0.371	0.402
60 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	171.0	238.9	267.4	291.2	313.6	335.4	367.8	401.4	435.8	470.3
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.390	0.517	0.568	0.607	0.640	0.671	0.716	0.756	0.795	0.833
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.051	0.082	0.121	0.150	0.174	0.195	0.224	0.250	0.275	0.301
80 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	146.4	205.0	248.3	277.6	302.8	326.4	360.5	395.3	430.8	466.1
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.340	0.477	0.522	0.569	0.608	0.641	0.688	0.731	0.770	0.808
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.044	0.057	0.082	0.109	0.131	0.149	0.174	0.197	0.219	0.239
100 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	143.5	182.5	228.3	263.0	291.3	317.1	353.3	389.6	426.0	462.0
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.327	0.401	0.480	0.535	0.579	0.614	0.665	0.709	0.751	0.789
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.039	0.046	0.056	0.067	0.081	0.094	0.111	0.128	0.143	0.159
150 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	139.1	167.7	199.8	235.0	267.2	296.5	336.5	376.0	414.8	452.6
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.308	0.361	0.417	0.474	0.524	0.564	0.619	0.668	0.712	0.752
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.037	0.042	0.047	0.054	0.062	0.070	0.085	0.097	0.108	0.119
200 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	138.3	163.8	191.1	221.5	252.3	281.7	323.4	364.7	404.9	444.2
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.295	0.342	0.392	0.440	0.487	0.529	0.585	0.636	0.682	0.724
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.035	0.039	0.043	0.049	0.054	0.059	0.069	0.078	0.088	0.099
250 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	138.2	161.8	187.5	215.1	243.9	272.3	314.1	355.8	397.1	437.5
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.286	0.331	0.376	0.419	0.462	0.503	0.561	0.612	0.659	0.702
v	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.034	0.038	0.041	0.045	0.049	0.054	0.061	0.069	0.077	0.083
300 h	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	137.9	161.4	186.3	212.1	238.7	265.9	307.7	349.4	391.2	432.0
s	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	0.277	0.322	0.364	0.404	0.446	0.484	0.541	0.593	0.640	0.684

* York and White, *Trans. Am. Inst. Chem. Engrs.*, **40**, 227 (1944).

Properties of Saturated Methane*

Temp., °F.	Abs. pressure, lb./sq. in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy B.t.u./(lb.)(°R.)	
		Liquid	Vapor	Liquid	Vapor	Liquid	Vapor
t	p	v_f	v_g	h_f	h_g	s_f	s_g
−280	4.90	0.03635	24.04	0	228.2	0	1.2699
−270	8.44	.03698	14.61	8.2	232.3	0.0423	1.2236
−260	13.80	.03766	9.31	16.6	236.4	.0823	1.1830
−250	21.71	.03839	6.13	25.0	240.3	.1201	1.1468
−240	32.4	.03915	4.24	33.3	243.9	.1578	1.1164
−230	46.4	.03999	3.04	42.0	247.3	.1962	1.0900
−220	64.5	.04092	2.23	50.6	250.2	.2333	1.0660
−210	87.6	.04193	1.67	59.5	252.8	.2693	1.0434
−200	115.7	.04306	1.281	68.8	254.8	.3062	1.0224
−190	150.0	.04431	0.990	78.2	256.2	.3419	1.0019
−180	191.5	.04575	.773	87.8	257.0	.3767	.9816
−170	240.0	.04745	.610	98.0	257.2	.4127	.9622
−160	297.0	.04944	.483	108.7	256.5	.4476	.9411
−150	364	.05197	.381	120.3	254.5	.4839	.9169
−140	440	.05224	.3008	133.2	251.2	.5214	.8905
−130	527	.05999	.2318	148.1	245.9	.5656	.8622
−120	627	.06961	.1613	171.8	231.4	.6329	.8083
−115.8	673	.0983	.0983	203.4	203.4	.7232	.7232

* Matthews and Hurd, *Trans. Am. Inst. Chem. Engrs.* **42**, 55 (1946).

Properties of Superheated Methane*

v, volume, cu. ft./lb.; *h*, enthalpy, B.t.u./lb.; *s*, entropy, B.t.u./(lb.)(°R.)
Parenthetic figures after pressures are saturation temperatures

Temp., °F.	10 lb./sq. in. abs. (−266.6°F.)			20 lb./sq. in. abs. (−251.8°F.)			30 lb./sq. in. abs. (−242°F.)			40 lb./sq. in. abs. (−234.3°F.)			60 lb./sq. in. abs. (−222.2°F.)			80 lb./sq. in. abs. (−213.0°F.)			100 lb./sq. in. abs. (−205.5°F.)		
t	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*
−260	12.98	237.0	1.2262	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....
−240	14.39	247.3	1.2750	7.04	245.8	1.1830	4.60	244.2	1.1273	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....
−220	15.78	257.5	1.3214	7.76	256.3	1.2313	5.09	254.9	1.1767	3.75	253.6	1.1365	2.421	251.1	1.0775	.....	.....	.....	.....	.....	.....
−200	17.15	267.7	1.3644	8.47	266.6	1.2752	5.57	265.4	1.2217	4.12	264.3	1.1826	2.678	262.1	1.1258	1.954	259.5	1.0832	1.518	256.9	1.0476
−180	18.52	277.9	1.4032	9.17	276.9	1.3148	6.06	275.9	1.2618	4.49	274.9	1.2237	2.934	273.0	1.1681	2.153	270.7	1.1269	1.684	268.4	1.0935
−160	19.87	287.9	1.4386	9.86	287.0	1.3507	6.53	286.1	1.2982	4.85	285.2	1.2607	3.184	283.5	1.2062	2.348	281.5	1.1662	1.847	279.6	1.1339
−140	21.28	297.9	1.4710	10.56	297.1	1.3836	6.99	296.3	1.3315	5.21	295.5	1.2944	3.429	293.9	1.2407	2.539	292.3	1.2015	2.002	290.6	1.1701
−120	22.62	308.0	1.5009	11.25	307.2	1.4138	7.45	306.4	1.3621	5.56	305.8	1.3252	3.670	304.4	1.2721	2.725	302.9	1.2336	2.155	301.4	1.2029
−115.8	22.91	310.0	1.5067	11.40	309.2	1.4197	7.56	308.3	1.3681	5.64	307.9	1.3312	3.72	306.5	1.2782	2.761	305.0	1.2398	2.183	303.6	1.2092
−100	23.97	318.1	1.5290	11.94	317.4	1.4422	7.91	316.7	1.3908	5.91	316.1	1.3541	3.91	314.9	1.3015	2.903	313.5	1.2635	2.301	312.2	1.2332
− 80	25.33	328.1	1.5561	12.61	327.6	1.4693	8.37	327.0	1.4182	6.25	326.4	1.3816	4.14	325.2	1.3293	3.080	324.1	1.2918	2.444	322.9	1.2619
− 60	26.69	338.2	1.5820	13.29	337.6	1.4953	8.83	337.0	1.4443	6.60	236.6	1.4079	4.37	335.6	1.3558	3.255	334.5	1.3186	2.588	333.4	1.2891
− 40	28.02	348.3	1.6066	13.97	347.8	1.5200	9.28	347.3	1.4691	6.94	346.9	1.4328	4.60	345.9	1.3811	3.432	345.0	1.3440	2.729	344.0	1.3148
− 20	29.36	358.5	1.6303	14.64	358.0	1.5437	9.74	357.5	1.4929	7.29	357.1	1.4567	4.83	356.3	1.4052	3.607	355.4	1.3683	2.871	354.5	1.3393
0	30.72	368.7	1.6531	15.32	368.3	1.5667	10.19	367.9	1.5159	7.63	367.5	1.4798	5.06	366.7	1.4284	3.78	365.9	1.3917	3.014	365.2	1.3628
20	32.06	378.9	1.6752	16.00	378.6	1.5888	10.64	378.2	1.5380	7.97	377.8	1.5020	5.30	377.0	1.4508	3.96	376.4	1.4142	3.155	375.7	1.3854
40	33.40	389.4	1.6964	16.66	389.0	1.6101	11.10	388.6	1.5594	8.31	388.2	1.5234	5.52	387.6	1.4723	4.13	387.0	1.4358	3.293	386.3	1.4072
60	34.73	399.9	1.7169	17.34	399.5	1.6306	11.54	399.2	1.5800	8.65	398.9	1.5440	5.75	398.3	1.4930	4.30	397.7	1.4566	3.431	397.0	1.4280
80	36.10	410.5	1.7370	18.02	410.2	1.6507	12.00	409.9	1.6002	8.99	409.6	1.5642	5.98	409.1	1.5133	4.47	408.4	1.4770	3.569	407.9	1.4485
100	37.44	421.4	1.7570	18.70	421.1	1.6707	12.44	420.8	1.6202	9.33	420.5	1.5843	6.21	420.1	1.5334	4.65	419.4	1.4971	3.71	418.9	1.4687
120	38.78	432.2	1.7763	19.37	431.9	1.6900	12.90	431.6	1.6396	9.66	431.4	1.6036	6.44	431.0	1.5529	4.82	430.3	1.5166	3.86	429.8	1.4882
140	40.12	443.2	1.7946	20.05	442.9	1.7084	13.35	442.7	1.6580	10.01	442.4	1.6221	6.66	442.0	1.5714	4.99	441.4	1.5351	3.98	440.9	1.5068
160	41.46	454.4	1.8127	20.73	454.2	1.7265	13.79	453.9	1.6760	10.34	453.7	1.6402	6.89	453.3	1.5895	5.16	452.7	1.5533	4.12	452.2	1.5250
180	42.80	465.7	1.8307	21.40	465.5	1.7445	14.25	465.2	1.6940	10.69	465.0	1.6582	7.12	464.6	1.6075	5.33	464.1	1.5714	4.26	463.7	1.5432
200	44.13	477.3	1.8484	22.07	477.1	1.7622	14.70	476.9	1.7118	11.03	476.7	1.6760	7.34	476.3	1.6254	5.50	475.8	1.5893	4.40	475.4	1.5611
220	45.47	488.9	1.8659	22.74	488.7	1.7798	15.16	488.5	1.7294	11.36	488.3	1.6936	7.56	487.9	1.6430	5.66	487.4	1.6069	4.52	487.1	1.5788
240	46.81	500.9	1.8829	23.41	500.7	1.7968	15.61	500.5	1.7464	11.70	500.3	1.7106	7.77	499.9	1.6600	5.84	499.5	1.6241	4.66	499.2	1.5960
260	48.15	512.9	1.8998	24.08	512.8	1.8137	16.06	512.6	1.7633	12.04	512.4	1.7276	8.02	512.0	1.6670	6.01	511.6	1.6411	4.80	511.3	1.6130
280	49.49	525.1	1.9166	24.75	525.0	1.8305	16.50	524.8	1.7801	12.37	524.6	1.7444	8.24	524.2	1.6938	6.17	523.9	1.6579	4.93	523.5	1.6299
300	50.83	537.6	1.9331	25.42	537.4	1.8470	16.94	537.2	1.7966	12.71	537.0	1.7609	8.46	536.7	1.7103	6.35	536.3	1.6744	5.07	536.0	1.6464
320	52.16	550.2	1.9493	26.09	550.0	1.8633	17.39	549.8	1.8129	13.04	549.6	1.7771	8.69	549.3	1.7266	6.52	549.0	1.6907	5.21	548.6	1.6627
340	53.50	563.0	1.9655	26.75	562.9	1.8795	17.83	562.7	1.8291	13.38	562.5	1.7934	8.91	562.2	1.7429	6.68	561.9	1.7070	5.35	561.5	1.6790
360	54.84	576.0	1.9815	27.43	575.9	1.8955	18.28	575.7	1.8451	13.72	575.5	1.8096	9.14	575.2	1.7589	6.86	574.9	1.7230	5.49	574.6	1.6951
380	56.18	589.2	1.9973	28.10	589.1	1.9113	18.73	588.9	1.8609	14.05	588.7	1.8252	9.36	588.5	1.7747	7.03	588.2	1.7389	5.61	587.8	1.7110
400	57.51	602.6	2.0132	28.76	602.5	1.9271	19.17	602.4	1.8768	14.38	602.2	1.8411	9.58	602.0	1.7906	7.19	601.7	1.7548	5.75	601.3	1.7269
420	58.85	616.1	2.0290	29.43	616.0	1.9429	19.62	615.8	1.8926	14.72	615.7	1.8569	9.81	615.5	1.8064	7.36	615.3	1.7706	5.89	614.9	1.7427
440	60.19	629.9	2.0446	30.10	629.8	1.9585	20.07	629.7	1.9082	15.05	629.6	1.8725	10.04	629.4	1.8221	7.53	629.1	1.7862	6.02	628.8	1.7583
460	61.53	643.9	2.0601	30.77	643.8	1.9741	20.51	643.7	1.9237	15.38	643.6	1.8880	10.26	643.4	1.8376	7.69	643.2	1.8018	6.15	642.9	1.7739
480	62.87	658.0	2.0755	31.44	657.9	1.9895	20.96	657.8	1.9391	15.72	657.7	1.9034	10.48	657.5	1.8530	7.87	657.4	1.8172	6.28	657.0	1.7893
500	64.20	672.4	2.0907	32.10	672.3	2.0047	21.40	672.3	1.9543	16.05	672.2	1.9186	10.70	672.0	1.8682	8.03	671.8	1.8325	6.42	671.5	1.8046

Temp., °F.	150 lb./sq. in. abs. (−190.0°F.)			200 lb./sq. in. abs. (−178.2°F.)			300 lb./sq. in. abs. (−159.5°F.)			500 lb./sq. in. abs. (−132.9°F.)			800 lb./sq. in. abs.			1000 lb./sq. in. abs.			1500 lb./sq. in. abs.		
t	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*	*v*	*h*	*s*
−180	1.052	262.4	1.0267	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....
−160	1.172	274.5	1.0716	0.830	269.4	1.0240	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....
−140	1.283	286.3	1.1106	.923	281.6	1.0655	0.553	271.1	0.9899	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....	.....
−120	1.391	297.7	1.1452	1.010	293.5	1.1016	.624	284.7	1.0327	0.3000	261.9	0.9197	.....	.....	.....	.....	.....	.....	.....	.....	.....
−115.8	1.414	299.9	1.1518	1.027	295.8	1.1085	.637	287.4	1.0409	.3142	265.9	0.9313	.....	.....	.....	.....	.....	.....	.....	.....	.....
−100	1.495	308.9	1.1769	1.092	305.2	1.1346	.687	297.5	1.0706	.3566	280.4	0.9715	0.1441	236.9	0.8141	.....	.....	.....	.....	.....	.....
− 80	1.597	319.9	1.2065	1.172	316.6	1.1654	.747	309.9	1.1040	.402	295.5	1.0139	.1969	267.9	.8992	0.1262	238.3	0.8118	.....	.....	.....
− 60	1.695	330.7	1.2344	1.247	327.9	1.1941	.802	321.9	1.1345	.443	309.3	1.0500	.2359	287.9	.9514	.1650	267.8	.8865	0.0870	223.5	0.7723
− 40	1.793	341.5	1.2606	1.324	339.0	1.2211	.856	333.6	1.1627	.481	322.5	1.0819	.2674	304.3	.9921	.1957	289.2	.9397	.1069	253.3	.8373
− 20	1.890	352.3	1.2856	1.399	350.0	1.2467	.909	345.1	1.1889	.518	335.2	1.1111	.2953	319.4	1.0276	.2212	307.6	.9809	.1270	276.6	.8860
0	1.989	363.0	1.3096	1.475	360.9	1.2711	.961	356.5	1.2145	.551	347.7	1.1386	.3202	333.9	1.0595	.2438	324.0	1.0165	.1453	297.8	.9292
20	2.084	373.6	1.3326	1.548	371.8	1.2944	1.011	367.9	1.2386	.584	359.9	1.1645	.3441	347.8	1.0885	.2647	339.3	1.0481	.1617	317.3	.9657
40	2.177	384.5	1.3546	1.620	382.8	1.3167	1.062	379.3	1.2615	.617	372.2	1.1891	.367	361.4	1.1151	.2848	354.0	1.0764	.1770	335.5	.9983
60	2.269	395.4	1.3757	1.691	393.8	1.3380	1.111	390.6	1.2834	.649	384.3	1.2123	.389	374.6	1.1399	.3031	368.2	1.1025	.1912	352.4	1.0276
80	2.364	406.3	1.3964	1.762	404.9	1.3589	1.161	402.1	1.3047	.681	396.3	1.2347	.411	387.5	1.1638	.3217	381.6	1.1273	.2051	367.9	1.0554
100	2.459	417.5	1.4168	1.835	416.1	1.3795	1.211	413.4	1.3258	.712	408.1	1.2565	.432	400.1	1.1869	.3396	395.0	1.1513	.2182	382.5	1.0818
120	2.554	428.6	1.4366	1.906	427.3	1.3994	1.259	424.7	1.3460	.742	419.9	1.2774	.453	412.4	1.2088	.3568	407.5	1.1741	.2306	396.5	1.1069
140	2.648	439.8	1.4552	1.976	438.6	1.4182	1.307	436.2	1.3652	.773	431.6	1.2969	.473	424.6	1.2294	.374	420.1	1.1954	.2429	410.0	1.1302
160	2.739	451.2	1.4736	2.046	450.0	1.4367	1.355	447.7	1.3839	.803	443.4	1.3157	.493	436.9	1.2493	.390	432.8	1.2160	.2548	423.3	1.1524
180	2.833	462.7	1.4919	2.116	461.6	1.4551	1.403	459.4	1.4025	.833	455.3	1.3347	.513	449.2	1.2689	.407	445.3	1.2363	.2664	436.4	1.1741
200	2.924	474.4	1.5099	2.187	473.4	1.4732	1.450	471.4	1.4208	.862	467.4	1.3531	.532	461.8	1.2881	.422	458.1	1.2560	.2780	449.6	1.1948
220	3.012	486.1	1.5276	2.253	485.2	1.4911	1.496	483.4	1.4388	.891	479.7	1.3712	.551	474.4	1.3068	.438	470.9	1.2753	.2892	462.7	1.2150
240	3.106	498.4	1.5448	2.323	497.4	1.5084	1.544	495.8	1.4562	.920	492.3	1.3888	.571	487.1	1.3250	.454	484.0	1.2938	.3002	476.0	1.2343
260	3.197	510.6	1.5619	2.392	509.7	1.5255	1.591	508.0	1.4735	.949	504.8	1.4063	.590	499.8	1.3429	.470	496.9	1.3120	.3112	489.2	1.2531
280	3.287	523.0	1.5788	2.461	522.1	1.5425	1.638	520.4	1.4907	.978	517.4	1.4236	.608	512.7	1.3607	.485	509.8	1.3300	.3221	502.6	1.2716
300	3.378	535.5	1.5954	2.531	534.7	1.5592	1.684	533.0	1.5074	1.007	530.2	1.4406	.627	525.7	1.3781	.501	522.9	1.3475	.3326	516.2	1.2896
320	3.459	548.2	1.6117	2.599	547.4	1.5756	1.730	545.8	1.5239	1.036	534.2	1.4573	.646	538.8	1.3951	.516	536.1	1.3646	.3429	529.8	1.3072
340	3.559	561.1	1.6281	2.667	560.3	1.5919	1.776	558.8	1.5403	1.064	556.3	1.4740	.664	552.1	1.4120	.530	549.4	1.3817	.3534	543.5	1.3247
360	3.66	574.2	1.6442	2.736	573.4	1.6080	1.823	571.9	1.5566	1.093	569.6	1.4904	.682	565.6	1.4287	.545	563.0	1.3985	.364	557.3	1.3419
380	3.75	587.5	1.6601	2.805	586.7	1.6239	1.869	585.3	1.5726	1.121	583.1	1.5066	.700	579.2	1.4450	.560	576.7	1.4149	.374	571.3	1.3589

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* Matthews and Hurd, *Trans. Am. Inst. Chem. Engrs.* **42**, 55 (1946).

Properties of Saturated Carbon Dioxide*†‡

Temp., °F.	Abs. pressure, lb./sq. in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy B.t.u./(lb.)(°R.)		Temp., °F.	Abs. pressure, lb./sq. in.	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy B.t.u./(lb.)(°R.)	
		Condensed phase*	Vapor	Condensed phase*	Vapor	Condensed phase*	Vapor			Condensed phase*	Vapor	Condensed phase*	Vapor	Condensed phase*	Vapor
t	p	v_f	v_g	h_f	h_g	s_f	s_g	t	p	v_f	v_g	h_f	h_g	s_f	s_g
−140	3.18	0.01008	24.320	−121.5	129.2	0.6065	1.3908	− 20	214.9	.01498	.4168	9.1	138.5	.9430	1.2372
−120	8.90	.01018	9.179	−116.0	132.0	.6232	1.3636	− 10	257.3	.01532	.3472	13.9	138.7	.9532	1.2303
−100	22.22	.01032	3.804	−110.1	134.3	.6403	1.3199	0	305.5	.01570	.2904	18.8	138.9	.9636	1.2247
− 90	33.98	.01040	2.525	−106.7	135.1	.6499	1.3033								
− 80	50.85	.01048	1.700	−102.5	135.7	.6607	1.2881	10	360.2	.01614	.2437	24.0	138.7	.9744	1.2188
								20	421.8	.01663	.2049	29.4	138.3	.9856	1.2127
− 70	74.82	.01059	1.162	− 98.0	135.9	.6724	1.2726	30	490.8	.01719	.1722	35.4	137.8	.9976	1.2067
− 69.9	75.10	.01059	1.157	− 97.9	135.9	.6725	1.2724	40	567.8	.01787	.1444	41.7	136.7	1.0092	1.1994
− 69.9	75.10	.01360	1.1570	− 13.7	135.9	.8885	1.2724	50	653.6	.01868	.1205	48.4	135.0	1.0218	1.1917
− 60	94.7	.01384	0.9270	− 9.2	136.6	.8997	1.2647								
− 50	118.2	.01409	.7492	− 4.7	137.2	.9110	1.2572	60	748.6	.01970	.0994	55.5	132.1	1.0353	1.1826
								70	853.4	.02112	.08040	63.7	127.5	1.0500	1.1724
− 40	145.8	.01437	.6113	.00	137.8	.9218	1.2503	80	968.7	.02370	.06064	73.9	118.7	1.0694	1.1555
− 30	177.8	.01466	.5029	4.5	138.2	.9325	1.2436	87.8	1069.4	.03454	.03454	97.0	97.0	1.1098	1.1098

* Above the solid line the condensed phase is solid; below the line it is liquid.

† "Refrigerating Data Book," 5th ed., American Society of Refrigerating Engineers, New York, 1942.

‡ s_f = 1.0 at 32°F.
h_f = 36.7 at 32°F.

Properties of Superheated Carbon Dioxide*,†

v, volume, cu. ft./lb.; h, enthalpy, B.t.u./lb.; s, entropy, B.t.u/(lb.)(°R.); p, absolute pressure, lb./sq. in.

p		−75°F.	−50°F.	0°F.	50°F.	100°F.	150°F.	200°F.	300°F.	400°F.	600°F.	800°F.	1000°F.	1200°F.	1400°F.	1600°F.	1800°F.
1.00	v	93.90	100.0	112.2	124.4	136.6	148.8	161.0	185.4	209.7	258.5	307.2	356.0	404.8	453.6	502.3	551.0
	h	283.2	288.0	297.8	307.7	318.0	328.4	339.1	361.4	384.7	434.4	487.1	542.4	599.6	658.6	718.8	780.0
	s	1.4772	1.4892	1.5112	1.5316	1.5506	1.5684	1.5852	1.6165	1.6451	1.6969	1.7423	1.7829	1.8197	1.8533	1.8838	1.9123
10.0	v	9.280	9.902	11.15	12.38	13.61	14.84	16.06	18.51	20.96	25.85	30.73	35.61	40.49	45.36	50.24	55.11
	h	282.6	287.5	297.3	307.3	317.7	328.2	339.0	361.3	384.6	434.4	487.1	542.4	599.6	658.6	718.8	780.0
	s	1.3733	1.3853	1.4073	1.4277	1.4467	1.4645	1.4813	1.5126	1.5412	1.5930	1.6384	1.6790	1.7158	1.7494	1.7799	1.8084
20.0	v	4.586	4.904	5.542	6.119	6.778	7.407	8.016	9.247	10.47	12.92	15.36	17.80	20.24	22.68	25.11	27.55
	h	281.9	287.0	296.8	306.8	317.3	327.9	338.8	361.1	384.5	434.3	487.1	542.4	599.6	658.6	718.8	780.0
	s	1.3417	1.3538	1.3759	1.3964	1.4154	1.4332	1.4500	1.4813	1.5099	1.5617	1.6071	1.6477	1.6845	1.7181	1.7486	1.7771
40.0	v	2.239	2.404	2.738	3.053	3.363	3.688	3.993	4.615	5.230	6.458	7.688	8.901	10.12	11.37	12.56	13.78
	h	280.6	285.9	295.8	305.9	316.5	327.4	338.4	360.9	384.3	434.2	487.0	542.4	599.6	658.6	718.8	780.0
	s	1.3088	1.3211	1.3435	1.3642	1.3834	1.4014	1.4184	1.4499	1.4787	1.5305	1.5759	1.6165	1.6533	1.6869	1.7174	1.7459
80.0	v	......	1.154	1.335	1.498	1.657	1.828	1.982	2.298	2.608	3.226	3.839	4.448	5.060	5.670	6.281	6.887
	h	......	283.8	293.8	304.1	315.1	326.4	337.7	360.2	383.9	434.0	486.9	542.3	599.5	658.6	718.8	780.0
	s	......	1.2778	1.3044	1.3284	1.3490	1.3679	1.3855	1.4177	1.4468	1.4991	1.5446	1.5852	1.6220	1.6556	1.6861	1.7146
120	v	......	......	0.8665	0.9799	1.088	1.208	1.311	1.525	1.734	2.148	2.559	2.966	3.373	3.781	4.188	4.592
	h	......	......	291.7	302.2	313.6	325.4	337.0	359.7	383.5	433.8	486.8	542.3	599.5	658.6	718.8	780.0
	s	......	......	1.2833	1.3086	1.3297	1.3488	1.3666	1.3993	1.4285	1.4808	1.5263	1.5669	1.6037	1.6373	1.6678	1.6963
160	v	......	......	0.6305	0.7207	0.8033	0.8986	0.9760	1.139	1.297	1.610	1.918	2.224	2.530	2.836	3.141	3.445
	h	......	......	289.7	300.4	312.1	324.4	336.3	359.1	383.1	433.6	486.6	542.2	599.5	658.6	718.8	780.0
	s	......	......	1.2666	1.2928	1.3154	1.3350	1.3529	1.3857	1.4151	1.4675	1.5133	1.5539	1.5907	1.6243	1.6548	1.6833
200	v	......	......	0.4891	0.5652	0.6376	0.7125	0.7748	0.9075	1.035	1.287	1.534	1.779	2.024	2.269	2.513	2.757
	h	......	......	287.7	298.6	310.6	323.4	335.6	358.5	382.7	433.4	486.5	542.2	599.5	658.5	718.8	780.0
	s	......	......	1.2519	1.2805	1.3038	1.3239	1.3421	1.3753	1.4049	1.4574	1.5033	1.5439	1.5807	1.6143	1.6448	1.6733
240	v	......	......	0.3948	0.4614	0.5237	0.5886	0.6407	0.7532	0.8604	1.071	1.273	1.482	1.687	1.891	2.095	2.297
	h	......	......	285.6	296.7	309.1	322.4	334.9	358.0	382.3	433.1	486.4	542.1	599.5	658.5	718.8	780.0
	s	......	......	1.2395	1.2694	1.2940	1.3145	1.3330	1.3671	1.3963	1.4490	1.4948	1.5356	1.5724	1.6060	1.6365	1.6650
300	v	......	......	......	0.3563	0.4100	0.4636	0.5065	0.5985	0.6868	0.8556	1.021	1.186	1.349	1.513	1.676	1.838
	h	......	......	......	294.0	306.9	320.9	333.9	357.1	381.6	432.8	486.2	542.0	599.4	658.5	718.7	780.0
	s	......	......	......	1.2562	1.2813	1.3029	1.3219	1.3560	1.3862	1.4389	1.4848	1.5256	1.5624	1.5960	1.6265	1.6550
360	v	......	......	......	0.2858	0.3341	0.3780	0.4171	0.4958	0.5693	0.7212	0.8502	0.9874	1.125	1.261	1.397	1.533
	h	......	......	......	291.2	304.6	319.4	332.8	356.3	381.0	432.5	486.0	541.9	599.4	658.5	718.7	779.9
	s	......	......	......	1.2436	1.2699	1.2925	1.3124	1.3475	1.3779	1.4307	1.4766	1.5174	1.5542	1.5878	1.6183	1.6468
440	v	......	......	......	0.2216	0.2652	0.3040	0.3358	0.4022	0.4633	0.5817	0.6950	0.8079	0.9201	1.032	1.142	1.255
	h	......	......	......	287.6	301.6	317.4	331.4	355.1	380.2	432.1	485.8	541.7	599.3	658.4	718.6	779.9
	s	......	......	......	1.2282	1.2559	1.2797	1.3006	1.3370	1.3681	1.4215	1.4675	1.5083	1.5451	1.5787	1.6092	1.6377
520	v	......	......	......	0.1772	0.2174	0.2513	0.2795	0.3374	0.3901	0.4912	0.5881	0.6832	0.7785	0.8733	0.9672	1.062
	h	......	......	......	283.9	298.7	315.4	330.0	354.0	379.4	431.7	485.5	541.5	599.2	658.3	718.6	779.9
	s	......	......	......	1.2148	1.2438	1.2687	1.2905	1.3281	1.3599	1.4138	1.4599	1.5007	1.5375	1.5711	1.6010	1.6301
600	v	......	......	......	0.1452	0.1823	0.2123	0.2383	0.2898	0.3363	0.4250	0.5093	0.5921	0.6747	0.7571	0.8385	0.9202
	h	......	......	......	280.3	295.7	313.4	328.6	352.8	378.6	431.1	485.3	541.4	599.0	658.2	718.6	779.8
	s	......	......	......	1.2020	1.2323	1.2583	1.2809	1.3198	1.3525	1.4071	1.4534	1.4942	1.5310	1.5646	1.5951	1.6236
800	v	......	......	......	......	0.1196	0.1483	0.1712	0.2126	0.2489	0.3173	0.3812	0.4436	0.5060	0.5680	0.6292	0.6906
	h	......	......	......	......	288.2	308.4	325.1	350.0	376.5	430.1	484.7	541.0	598.8	658.0	718.4	779.7
	s	......	......	......	......	1.2111	1.2391	1.2631	1.3041	1.3380	1.3935	1.4404	1.4812	1.5180	1.5516	1.5821	1.6106
1000	v	......	......	......	......	......	0.1101	0.1310	0.1663	0.1966	0.2526	0.3048	0.3547	0.4049	0.4545	0.5037	0.5526
	h	......	......	......	......	......	303.4	321.6	347.1	374.5	429.1	484.0	540.6	598.5	657.8	718.3	779.6
	s	......	......	......	......	......	1.2218	1.2472	1.2903	1.3258	1.3828	1.4302	1.4712	1.5080	1.5416	1.5721	1.6006
1200	v	......	......	......	......	......	......	0.1042	0.1356	0.1621	0.2096	0.2531	0.2953	0.3374	0.3789	0.4199	0.4609
	h	......	......	......	......	......	......	318.4	344.2	372.5	428.1	483.5	540.2	598.2	657.7	718.2	779.5
	s	......	......	......	......	......	......	1.2343	1.2791	1.3158	1.3740	1.4216	1.4628	1.4996	1.5332	1.5637	1.5922
1400	v	......	......	......	......	......	......	......	0.1136	0.1375	0.1788	0.2160	0.2529	0.2892	0.3249	0.3601	0.3551
	h	......	......	......	......	......	......	......	341.4	370.4	427.0	482.9	539.8	598.0	657.5	718.0	779.5
	s	......	......	......	......	......	......	......	1.2703	1.3078	1.3668	1.4145	1.4558	1.4927	1.5263	1.5568	1.5853
1600	v	......	......	......	......	......	......	......	......	0.1191	0.1557	0.1898	0.2211	0.2530	0.2843	0.3153	0.3461
	h	......	......	......	......	......	......	......	......	367.6	426.0	482.3	539.5	597.7	657.2	717.9	779.4
	s	......	......	......	......	......	......	......	......	1.3002	1.3602	1.4083	1.4497	1.4867	1.5193	1.5508	1.5793
1800	v	......	......	......	......	......	......	......	......	0.1047	0.1377	0.1675	0.1964	0.2249	0.2528	0.2804	0.3079
	h	......	......	......	......	......	......	......	......	364.0	424.9	481.7	539.1	597.4	657.0	717.8	779.3
	s	......	......	......	......	......	......	......	......	1.2930	1.3539	1.4023	1.4440	1.4812	1.5148	1.5453	1.5738

* Sweigert, Weber, and Allen, *Ind. Eng. Chem.*, **38**, 185 (1946); with permission.

† s_f = 1.0 at 32°F. h_f = 180 at 32°F. Therefore, according to the bases of Table 208, the entropies of these two CO_2 tables are consistent, but (180 − 36.7) or 143.3 B.t.u./lb. must be added to the enthalpies of saturated CO_2 to make them consistent with those of superheated CO_2.

Properties of Saturated Steam: Pressure Table*

Abs. pressure lb./sq. in. p	Temp., °F. t	Volume, cu. ft./lb.		Enthalpy, B.t.u./lb.		Entropy, B.t.u./(lb.)(°R.)		Internal energy, B.t.u./lb.	
		Liquid v_f	Vapor v_g	Liquid h_f	Vapor h_g	Liquid s_f	Vapor s_g	Liquid u_f	Vapor u_g
1.0	101.74	0.01614	333.6	69.70	1106.0	0.1326	1.9782	69.70	1044.3
2.0	126.08	.01623	173.73	93.99	1116.3	.1749	1.9200	93.98	1051.9
3.0	141.48	.01630	118.71	109.37	1122.6	.2008	1.8863	109.36	1056.7
4.0	152.97	.01636	90.63	120.86	1127.3	.2198	1.8625	120.85	1060.2
5.0	162.24	.01640	73.52	130.13	1131.1	.2347	1.8441	130.12	1063.1
6.0	170.06	.01645	61.98	137.96	1134.2	.2472	1.8292	137.94	1065.4
7.0	176.85	.01649	53.64	144.76	1136.9	.2581	1.8167	144.74	1067.4
8.0	182.86	.01653	47.34	150.79	1139.3	.2674	1.8057	150.77	1069.2
9.0	188.28	.01656	42.40	156.22	1141.4	.2759	1.7962	156.19	1070.8
10	193.21	.01659	38.42	161.17	1143.3	.2835	1.7876	161.14	1072.2
14.696	212.00	.01672	26.80	180.07	1150.4	.3120	1.7566	180.02	1077.5
15	213.03	.01672	26.29	181.11	1150.8	.3135	1.7549	181.06	1077.8
20	227.96	.01683	20.089	196.16	1156.3	.3356	1.7319	196.10	1081.9
25	240.07	.01692	16.303	208.42	1160.6	.3533	1.7139	208.34	1085.1
30	250.33	.01701	13.746	218.82	1164.1	.3680	1.6993	218.73	1087.8
35	259.28	.01708	11.898	227.91	1167.1	.3807	1.6870	227.80	1090.1
40	267.25	.01715	10.498	236.03	1169.7	.3919	1.6763	235.90	1092.0
45	274.44	.01721	9.401	243.36	1172.0	.4019	1.6669	243.22	1093.7
50	281.01	.01727	8.515	250.09	1174.1	.4110	1.6585	249.93	1095.3
55	287.07	.01732	7.787	256.30	1175.9	.4193	1.6509	256.12	1096.7
60	292.71	.01738	7.175	262.09	1177.6	.4270	1.6438	261.90	1097.9
65	297.97	.01743	6.655	267.50	1179.1	.4342	1.6374	267.29	1099.1
70	302.92	.01748	6.206	272.61	1180.6	.4409	1.6315	272.38	1100.2
75	307.60	.01753	5.816	277.43	1181.9	.4472	1.6259	277.19	1101.2
80	312.03	.01757	5.472	282.02	1183.1	.4531	1.6207	281.76	1102.1
85	316.25	.01761	5.168	286.39	1184.2	.4587	1.6158	286.11	1102.9
90	320.27	.01766	4.896	290.56	1185.3	.4641	1.6112	290.27	1103.7
95	324.12	.01770	4.652	294.56	1186.2	.4692	1.6068	294.25	1104.5
100	327.81	.01774	4.432	298.40	1187.2	.4740	1.6026	298.08	1105.2
110	334.77	.01782	4.049	305.66	1188.9	.4832	1.5948	305.30	1106.5
120	341.25	.01789	3.728	312.44	1190.4	.4916	1.5878	312.05	1107.6
130	347.32	.01796	3.455	318.81	1191.7	.4995	1.5812	318.38	1108.6
140	353.02	.01802	3.220	324.82	1193.0	.5069	1.5751	324.35	1109.6
150	358.42	.01809	3.015	330.51	1194.1	.5138	1.5694	330.01	1110.5
160	363.53	.01815	2.834	335.93	1195.1	.5204	1.5640	335.39	1111.2
170	368.41	.01822	2.675	341.09	1196.0	.5266	1.5590	340.52	1111.9
180	373.06	.01827	2.532	346.03	1196.9	.5325	1.5542	345.42	1112.5
190	377.51	.01833	2.404	350.79	1197.6	.5381	1.5497	350.15	1113.1
200	381.79	.01839	2.288	355.36	1198.4	.5435	1.5453	354.68	1113.7
250	400.95	.01865	1.8438	376.00	1201.1	.5675	1.5263	375.14	1115.8
300	417.33	.01890	1.5433	393.84	1202.8	.5879	1.5104	392.79	1117.1
350	431.72	.01913	1.3260	409.69	1203.9	.6056	1.4966	408.45	1118.0
400	444.59	.0193	1.1613	424.0	1204.5	.6214	1.4844	422.6	1118.5
450	456.28	.0195	1.0320	437.2	1204.6	.6356	1.4734	435.5	1118.7
500	467.01	.0197	0.9278	499.4	1204.4	.6487	1.4634	447.6	1118.6
550	476.94	.0199	.8424	460.8	1203.9	.6608	1.4542	458.8	1118.2
600	486.21	.0201	.7698	471.6	1203.2	.6720	1.4454	469.4	1117.7
650	494.90	.0203	.7083	481.8	1202.3	.6826	1.4374	479.4	1117.1
700	503.10	.0205	.6554	491.5	1201.2	.6925	1.4296	488.8	1116.3
750	510.86	.0207	.6092	500.8	1200.0	.7019	1.4223	598.0	1115.4
800	518.23	.0209	.5687	509.7	1198.6	.7108	1.4153	506.6	1114.4
850	525.26	.0210	.5327	518.3	1197.1	.7194	1.4085	515.0	1113.3
900	531.98	.0212	.5006	526.6	1195.4	.7275	1.4020	523.1	1112.1
950	538.43	.0214	.4717	534.6	1193.7	.7355	1.3957	530.9	1110.8
1000	544.61	.0216	.4456	542.4	1191.8	.7430	1.3897	538.4	1109.4
1100	556.31	.0220	.4001	557.4	1187.8	.7575	1.3780	552.9	1106.4
1200	567.22	.0223	.3619	571.7	1183.4	.7711	1.3667	566.7	1103.0
1300	577.46	.0227	.3293	585.4	1178.6	.7840	1.3559	580.0	1099.4
1400	587.10	.0231	.3012	598.7	1173.4	.7963	1.3454	592.7	1095.4
1500	596.23	.0235	.2765	611.6	1167.9	.8082	1.3351	605.1	1091.2
2000	635.82	.0257	.1878	671.7	1135.1	.8619	1.2849	662.2	1065.6
2500	668.13	.0287	.1307	730.6	1091.1	.9126	1.2322	717.3	1030.6
3000	695.36	.0346	.0858	802.5	1020.3	.9731	1.1615	783.4	972.7
3206.2	705.40	.0503	.0503	902.7	902.7	1.0580	1.0580	872.9	872.9

* Abridged from Keenan and Keyes, "Thermodynamic Properties of Steam," Wiley, New York, 1936. Copyright, 1937, by Joseph H. Keenan and Frederick G. Keyes.

Properties of Superheated Steam*

v, volume, cu. ft./lb.; *h*, enthalpy, B.t.u./lb.; *s*, entropy, B.t.u./(lb.)(°R.)

Abs. pressure, lb./sq. in. (sat. temp.)		Temp., °F.												
		200	300	400	500	600	700	800	900	1000	1100	1200	1400	1600
	v	392.6	452.3	512.0	571.6	631.2	690.8	750.4	809.9	869.5	929.1	988.7	1107.8	1227.0
1	h	1150.4	1195.8	1241.7	1288.3	1335.7	1383.8	1432.8	1482.7	1533.5	1585.2	1637.7	1745.7	1857.5
(101.74)	s	2.0512	2.1153	2.1720	2.2233	2.2702	2.3137	2.3542	2.3923	2.4283	2.4625	2.4952	2.5566	2.6137
	v	78.16	90.25	102.26	114.22	126.16	138.10	150.03	161.95	173.87	185.79	197.71	221.6	245.4
5	h	1148.8	1195.0	1241.2	1288.0	1335.4	1383.6	1432.7	1482.6	1533.4	1585.1	1637.7	1745.7	1857.4
(162.24)	s	1.8718	1.9370	1.9942	2.0456	2.0927	2.1361	2.1767	2.2148	2.2509	2.2851	2.3178	2.3792	2.4363
	v	38.85	45.00	51.04	57.05	63.03	69.01	74.98	80.95	86.92	92.88	98.84	110.77	122.69
10	h	1146.6	1193.9	1240.6	1287.5	1335.1	1383.4	1432.5	1482.4	1533.2	1585.0	1637.6	1745.6	1857.3
(193.21)	s	1.7927	1.8595	1.9172	1.9689	2.0160	2.0596	2.1002	2.1383	2.1744	2.2086	2.2413	2.3028	2.3598
	v	......	30.53	34.68	38.78	42.86	46.94	51.00	55.07	59.13	63.19	67.25	75.37	83.48
14.696	h	......	1192.8	1239.9	1287.1	1334.8	1383.2	1432.3	1482.3	1533.1	1584.8	1637.5	1745.5	1857.3
(212.00)	s	......	1.8160	1.8743	1.9261	1.9734	2.0170	2.0576	2.0958	2.1319	2.1662	2.1989	2.2603	2.3174
	v	......	22.36	25.43	28.46	31.47	34.47	37.46	40.45	43.44	46.42	49.41	55.37	61.34
20	h	......	1191.6	1239.2	1286.6	1334.4	1382.9	1432.1	1482.1	1533.0	1584.7	1637.4	1745.4	1857.2
(227.96)	s	......	1.7808	1.8396	1.8918	1.9392	1.9829	2.0235	2.0618	2.0978	2.1321	2.1648	2.2263	2.2834
	v	......	11.040	12.628	14.168	15.688	17.198	18.702	20.20	21.70	23.20	24.69	27.68	30.66
40	h	......	1186.8	1236.5	1284.8	1333.1	1381.9	1431.3	1481.4	1532.4	1584.3	1637.0	1745.1	1857.0
(267.25)	s	......	1.6994	1.7608	1.8140	1.8619	1.9058	1.9467	1.9850	2.0212	2.0555	2.0883	2.1498	2.2069
	v	......	7.259	8.357	9.403	10.427	11.441	12.449	13.452	14.454	15.453	16.451	18.446	20.44
60	h	......	1181.6	1233.6	1283.0	1331.8	1380.9	1430.5	1480.8	1531.9	1583.8	1636.6	1744.8	1856.7
(292.71)	s	......	1.6492	1.7135	1.7678	1.8162	1.8605	1.9015	1.9400	1.9762	2.0106	2.0434	2.1049	2.1621
	v	......	......	6.220	7.020	7.797	8.562	9.322	10.077	10.830	11.582	12.332	13.830	15.325
80	h	......	......	1230.7	1281.1	1330.5	1379.9	1429.7	1480.1	1531.3	1583.4	1636.2	1744.5	1856.5
(312.03)	s	......	......	1.6791	1.7346	1.7836	1.8281	1.8694	1.9079	1.9442	1.9787	2.0115	2.0721	2.1303
	v	......	......	4.937	5.589	6.218	6.835	7.446	8.052	8.656	9.259	9.860	11.060	12.258
100	h	......	......	1227.6	1279.1	1329.1	1378.9	1428.9	1479.5	1530.8	1582.9	1635.7	1744.2	1856.2
(327.81)	s	......	......	1.6518	1.7085	1.7581	1.8029	1.8443	1.8829	1.9193	1.9538	1.9867	2.0484	2.1056
	v	......	......	4.081	4.636	5.165	5.683	6.195	6.702	7.207	7.710	8.212	9.214	10.213
120	h	......	......	1224.4	1277.2	1327.7	1377.8	1428.1	1478.8	1530.2	1582.4	1635.3	1743.9	1856.0
(341.25)	s	......	......	1.6287	1.6869	1.7370	1.7822	1.8237	1.8625	1.8990	1.9335	1.9664	2.0281	2.0854
	v	......	......	3.468	3.954	4.413	4.861	5.301	5.758	6.172	6.604	7.035	7.895	8.752
140	h	......	......	1221.1	1275.2	1326.4	1376.8	1427.3	1478.2	1529.7	1581.9	1634.9	1743.5	1855.7
(353.02)	s	......	......	1.6087	1.6683	1.7190	1.7645	1.8063	1.8451	1.8817	1.9163	1.9493	2.0110	2.0683
	v	......	......	3.008	3.443	3.849	4.244	4.631	5.015	5.396	5.775	6.152	6.906	7.656
160	h	......	......	1217.6	1273.1	1325.0	1375.7	1426.4	1477.5	1529.1	1581.4	1634.5	1743.2	1855.5
(363.53)	s	......	......	1.5908	1.6519	1.7033	1.7491	1.7911	1.8301	1.8667	1.9014	1.9344	1.9962	2.0535
	v	......	......	2.649	3.044	3.411	3.764	4.110	4.452	4.792	5.129	5.466	6.136	6.804
180	h	......	......	1214.0	1271.0	1323.5	1374.7	1425.6	1476.8	1528.6	1581.0	1634.1	1742.9	1855.2
(373.06)	s	......	......	1.5745	1.6373	1.6894	1.7355	1.7776	1.8167	1.8534	1.8882	1.9212	1.9831	2.0404
	v	......	......	2.361	2.726	3.060	3.380	3.693	4.002	4.309	4.613	4.917	5.521	6.123
200	h	......	......	1210.3	1268.9	1322.1	1373.6	1424.8	1476.2	1528.0	1580.5	1633.7	1742.6	1855.0
(381.79)	s	......	......	1.5594	1.6240	1.6767	1.7232	1.7655	1.8048	1.8415	1.8763	1.9094	1.9713	2.0287
	v	......	......	2.125	2.465	2.772	3.066	3.352	3.634	3.913	4.191	4.467	5.017	5.565
220	h	......	......	1206.5	1266.7	1320.7	1372.6	1424.0	1475.5	1527.5	1580.0	1633.3	1742.3	1854.7
(389.86)	s	......	......	1.5453	1.6117	1.6652	1.7120	1.7545	1.7939	1.8308	1.8656	1.8987	1.9607	2.0181
	v	......	......	1.9276	2.247	2.533	2.804	3.068	3.327	3.584	3.839	4.093	4.597	5.100
240	h	......	......	1202.5	1264.5	1319.2	1371.5	1423.2	1474.8	1526.9	1579.6	1632.9	1742.0	1854.5
(397.37)	s	......	......	1.5319	1.6003	1.6546	1.7017	1.7444	1.7839	1.8209	1.8558	1.8889	1.9510	2.0084
	v	......	......	......	2.063	2.330	2.582	2.827	3.067	3.305	3.541	3.776	4.242	4.707
260	h	......	......	......	1262.3	1317.7	1370.4	1422.3	1474.2	1526.3	1579.1	1632.5	1741.7	1854.2
(404.42)	s	......	......	......	1.5892	1.6447	1.6922	1.7352	1.7748	1.8118	1.8467	1.8799	1.9420	1.9995
	v	......	......	......	1.9047	2.156	2.392	2.621	2.845	3.066	3.286	3.504	3.938	4.370
280	h	......	......	......	1260.0	1316.2	1369.4	1421.5	1473.5	1525.8	1578.6	1632.1	1741.4	1854.0
(411.05)	s	......	......	......	1.5796	1.6354	1.6834	1.7265	1.7662	1.8033	1.8383	1.8716	1.9337	1.9912
	v	......	......	......	1.7675	2.005	2.227	2.442	2.652	2.859	3.065	3.269	3.674	4.078
300	h	......	......	......	1257.6	1314.7	1368.3	1420.6	1472.8	1525.2	1578.1	1631.7	1741.0	1853.7
(417.33)	s	......	......	......	1.5701	1.6268	1.6751	1.7184	1.7582	1.7954	1.8305	1.8638	1.9260	1.9835
	v	......	......	......	1.4923	1.7036	1.8980	2.084	2.266	2.445	2.622	2.798	3.147	3.493
350	h	......	......	......	1251.5	1310.9	1365.5	1418.5	1471.1	1523.8	1577.0	1630.7	1740.3	1853.1
(431.72)	s	......	......	......	1.5481	1.6070	1.6563	1.7002	1.7403	1.7777	1.8130	1.8463	1.9086	1.9663
	v	......	......	......	1.2851	1.4770	1.6508	1.8161	1.9767	2.134	2.290	2.445	2.751	3.055
400	h	......	......	......	1245.1	1306.9	1362.7	1416.4	1469.4	1522.4	1575.8	1629.6	1739.5	1852.5
(444.59)	s	......	......	......	1.5281	1.5894	1.6398	1.6842	1.7247	1.7623	1.7977	1.8311	1.8936	1.9513

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* Abridged from Keenan and Keyes, "Thermodynamic Properties of Steam," Wiley, New York, 1936. Copyright, 1937, by Joseph H. Keenan and Frederick G. Keyes.

NOTES:

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