Chapter 3
Gas and Oil Properties and Correlations

3.1 Introduction
Chap. 3 covers the properties of oil and gas systems, their nomenclature and units, and correlations used for their prediction. Sec. 3.2 covers the fundamental engineering quantities used to describe phase behavior, including molecular quantities, critical and reduced properties, component fractions, mixing rules, volumetric properties, transport properties, and interfacial tension (IFT).

Sec. 3.3 discusses the properties of gas mixtures, including correlations for Z factor, pseudocrITICAL properties and wellstream gravity, gas viscosity, dewpoint pressure, and total formation volume factor (PVF). Sec. 3.4 covers oil properties, including correlations for bubblepoint pressure, compressibility, PVF, density, and viscosity. Sec. 3.5 gives correlations for IFT and diffusion coefficients. Sec. 3.6 reviews the estimation of K values for low-pressure applications, such as surface separator design, and convergence-pressure methods used for reservoir calculations.

3.2 Review of Properties, Nomenclature, and Units
3.2.1 Molecular Quantities. All matter is composed of elements that cannot be decomposed by ordinary chemical reactions. Carbon (C), hydrogen (H), sulfur (S), nitrogen (N), and oxygen (O) are examples of the elements found in naturally occurring petroleum systems.

The physical unit of the element is the atom. Two or more elements may combine to form a chemical compound. Carbon dioxide (CO₂), methane (CH₄), and hydrogen sulfide (H₂S) are examples of compounds found in naturally occurring petroleum systems. When two atoms of the same element combine, they form diatomic compounds, such as nitrogen (N₂) and oxygen (O₂). The physical unit of the compound is the mole.

Mass is the basic quantity for measuring the amount of a substance. Because chemical compounds always combine in a definite proportion (i.e., as a simple ratio of whole numbers), the mass of the atoms of different elements can be conveniently compared by relating them with a standard. The current standard is carbon-12, where the element carbon has been assigned a relative atomic mass of 12.011.

The relative atomic mass of all other elements have been determined relative to the carbon-12 standard. The smallest element is hydrogen, which has a relative atomic mass of 1.0079. The relative atomic mass of one element contains the same number of atoms as the relative atomic mass of any other element. This is true regardless of the units used to measure mass.

According to the SI standard, the definition of the mole reads “the mole is the amount of substance of a system which contains as many elementary entities as there are atoms in 0.012 kilograms of carbon-12.” The SI symbol for mole is mol, which is numerically identical to the traditional g mol.

The SPE/SI standard uses kmol as the unit for a mole where kmol designates “an amount of substance which contains as many kilograms (groups of molecules) as there are atoms in 12.0 kg (incorrectly written as 0.012 kg in the original SPE publication) of carbon-12 multiplied by the relative molecular mass of the substance involved.”

A practical way to interpret kmol is “kg mol” where kmol is numerically equivalent to 1,000 g mol (i.e., 1,000 mol). Otherwise, the following conversions apply.

1 kmol = 1,000 mol
= 1,000 g mol
= 2.2046 lbm mol

1 lbm mol = 0.45359 kmol
= 453.59 mol
= 453.59 g mol

1 mol = 1 g mol
= 0.001 kmol
= 0.0022046 lbm mol

The term molecular weight has been replaced in the SI system by molar mass. Molar mass, M, is defined as the mass per mole (M = m/n) of a given substance where the unit mole must be consistent with the unit of mass. The numerical value of molecular weight is independent of the units used for mass and moles, as long as the units are consistent. For example, the molar mass of methane is 16.04, which for various units can be written

\[ M = 16.04 \text{ kg/kmol} = 16.04 \text{ lbm/lbm mol} = 16.04 \text{ g/g mol} = 16.04 \text{ g/mol} \]

3.2.2 Critical and Reduced Properties. Most equations of state (EOS’s) do not use pressure and temperature explicitly to define the state of a system, but instead they generalize according to corresponding-states theory by use of two or more reduced properties, which are dimensionless.

\[ T_r = \frac{T}{T_c}, \quad \frac{P}{P_c}, \quad \frac{V}{V_c}, \quad \frac{Z}{Z_c} \]

(3.1a)
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The following relation for volume fractions \( x_i \) based on component densities at standard conditions \( \rho_i \) or specific gravities \( \gamma_i \):

\[
x_i = \frac{m_i/p_i}{\sum_{j=1}^{N} m_j/p_j}
\]

and \( \rho_i = \rho/p_c \)

\[
V_i = V/V_c
\]

where \( \rho = 1/V \). Absolute units must be used when calculating reduced pressure and temperature. \( \rho_c, T_c, V_c \), and \( \rho_v \) are the true critical properties of a pure component, or some average for a mixture. In most petroleum engineering applications, the range of reduced pressure is from 0.02 to 30 for gases and 0.03 to 40 for oils; reduced temperature ranges from <1 to 2.5 for gases and from 0.4 to 1.1 for oils. Reduced density can vary from 0 at low pressures to about 3.5 at high pressures.

Average mixture, or pseudocritical, properties are calculated from simple mixing rules or mixture specific gravity. Denoting a mixture pseudocritical property by \( \theta_{pc} \), the pseudocritical property is defined by \( \theta_{pc} = \theta/\theta_c \). Pseudocritical properties are not approximations of the true critical properties, but are chosen instead so that mixture properties will be estimated correctly with corresponding states correlations.

3.2.3 Component Fractions and Mixing Rules. Petroleum reservoir mixtures contain hundreds of well-defined and “undefined” components. These components are quantified on the basis of mole, weight, and volume fractions. For a mixture having \( N \) components, \( i = 1, \ldots, N \), the overall mole fractions are given by

\[
z_i = \frac{n_i}{N} = \frac{m_i/M}{\sum_{j=1}^{N} n_j/\sum_{j=1}^{N} m_j/\sum_{j=1}^{N} m_j/M_j}
\]

where \( n_i \) is the number of moles, \( m_i \) is mass, \( M \) is molecular weight, and the sum of \( z_i \) is 1.0. In general, oil composition is denoted by \( x_i \) and gas composition by \( y_i \).

Weight or mass fractions, \( w_i \), are given by

\[
w_i = \frac{m_i}{\sum_{j=1}^{N} m_j/\sum_{j=1}^{N} m_j/M_j}
\]

where the sum of \( w_i \) is 1.0. Although the composition of a mixture is usually expressed in terms of mole fraction, the measurement of composition is usually based on mass, which is converted to mole fraction with component molecular weights.

For oil mixtures at standard conditions (14.7 psia and 60°F), the total volume can be approximated by the sum of the volumes of individual components, assuming ideal-solution mixing. This results in the following relation for volume fractions \( x_i \):

\[
x_i = \frac{m_i/p_i}{\sum_{j=1}^{N} m_j/p_j}
\]

This mixing rule is usually adequate for molecular weight, pseudocritical temperature, andacentric factor. We can write a generalized linear mixing rule as

\[
\theta_{pc} = \frac{\sum_{i=1}^{N} \phi_i \theta_{i}}{\sum_{i=1}^{N} \phi_i}
\]

where \( \phi_i \) is usually one of the following weighting factors: \( \phi_i = z_i \) mole fraction (Kay’s rule); \( \phi_i = w_i \) weight fraction; or \( \phi_i = x_i \) volume fraction. Depending on the quantity being averaged, other mixing rules and definitions of \( \phi_i \) may be appropriate.

For example, the mixing rules used for constants in an EOS (Chap. 4) can be chosen on the basis of statistical thermodynamics.

3.2.4 Volumetric Properties. Density, \( \rho \), is defined as the ratio of mass to volume,

\[
\rho = m/V
\]

expressed in such units as lbm/ft³, kg/m³, and g/cm³. Fig. 3.1 shows the magnitudes of density for reservoir mixtures. Molar density, \( \rho_M \). Gives the volume per mole,

\[
\rho_M = n/V
\]

Specific volume, \( \tilde{v} \), is defined as the ratio of volume to mass and is equal to the reciprocal of density,

\[
\tilde{v} = V/m = 1/\rho
\]

Molar volume, \( v_M \), defines the ratio of volume per mole,

\[
v_M = n/V = M/\rho = 1/\rho_M
\]

and is typically used in cubic EOS’s. Molar density, \( \rho_M \), is given by

\[
\rho_M = 1/v = \rho/M
\]

and is used in the formulas of some EOS’s.

According to the SI standard, relative density replaces specific gravity as the term used to define the ratio of the density of a mixture to the density of a reference material. The conditions of pressure and temperature must be specified for both materials, and the densities of both materials are generally measured at standard conditions (standard conditions are usually 14.7 psia and 60°F).

\[
\gamma = \frac{\rho(p,p,T)}{\rho_{ref}(p_{ref},T_{ref})}
\]

Fig. 3.1—Reservoir densities as functions of pressure and temperature.
3.2.5 Black-Oil Pressure/Volume/Temperature (PVT) Properties.

The PVF, or simply volume factor, is used to convert a volume at elevated pressure and temperature to surface volume, and vice versa. More specifically, PVF is defined as the volume of a mixture at specified pressure and temperature divided by the volume of a product phase measured at standard conditions,

$$ B = \frac{V_{\text{surface}}}{V_{\text{product}}} \frac{p_T}{\rho_{\text{TO}}} $$

where $p_T$ is the total pressure in psia or psi. The units of $B$ are $\text{bbl/STB}$ for oil and water, and $\text{ft}^3/\text{scf}$ or $\text{bbl/Mscf}$ for gas. The surface product phase may consist of all or only part of the original mixture.

Primarily, four volume factors are used in petroleum engineering. They are oil PVF, $B_o$, water PVF, $B_w$, gas PVF, $B_g$, and total PVF of a gas/oil system, $B_t$, where

$$ B_o = \frac{V_o}{V_{\text{stock-tank-oil}}} $$

and

$$ B_g = \frac{V_g}{V_{\text{stock-tank-gas}}} $$

and

$$ B_w = \frac{V_w}{V_{\text{water/brine}}} $$

and the total PVF of a gas/water system is

$$ B_t = \frac{V_o + V_g + V_w}{V_{w/brine}} $$

In Eq. 3.17, $V_o$ is oil volume at $p$ and $T$; $V_g$ is gas volume at $p$ and $T$; $V_w$ = water/brine volume at $p$ and $T$; $V_{\text{stock-tank-oil}} = V_o$; $V_{\text{stock-tank-gas}} = V_g$; and $V_{w/brine} = V_w$ are reservoir volumes at standard conditions; and $\rho_{\text{TO}}$ is the density of the reservoir fluid.

Because gas PVF is inversely proportional to pressure, a reciprocal gas volume factor, $b_g$ (equal to $1/B_g$), is sometimes used, where the units of $b_g$ may be $\text{scf/ft}^3$ or $\text{Mscf/bbl}$. Fig. 3.3 shows PVF’s of typical reservoir systems. Inverse oil PVF, $b_o$, (equal to $1/B_o$) is also used in reservoir simulation.

Wet gas and gas-condensate reservoir fluids produce liquids at the surface, and for these gases the surface product (separator gas) consists of only part of the original reservoir gas mixture. Two gas PVF’s are used for these systems: the “dry” PVF, $B_d$, and the “wet” PVF, $B_w$, (or just $B_w$). $B_w$ gives the ratio of reservoir gas volume to the actual surface separator gas. $B_d$ gives the ratio of reservoir gas volume to a hypothetical “wet” surface-gas volume (the actual separation gas volume plus the stock-tank condensate converted to an equivalent surface-gas volume). Chap. 7 describes when $B_d$ and $B_w$ are used. The standard definition of $B_w$ is $\frac{p_T}{\rho_{\text{TO}}} \frac{(T)/p}{(T)/p}$ (see Eq. 3.38) represents the wet-gas PVF.
When a reservoir mixture produces both surface gas and oil, theGOR, \( R_{go} \), defines the ratio of standard gas volume to a reference oil volume (at stock-tank- or separator-oil volume),

\[
R_{go} = \frac{V_g}{V_o} = \frac{V_{g,sp}}{V_{o,sp}} \tag{3.18a}
\]

and \( R_{po} = \frac{V_p}{V_o} = \frac{V_{p,sp}}{V_{o,sp}} \tag{3.18b} \)
in units of scf/STB and scf/bbl, respectively. The separator condition should be reported when separator GOR is used.

Solution gas/oil ratio, \( R_s \), is the volume of gas (at standard conditions) liberated from a single-phase oil at elevated pressure and temperature divided by the resulting stock-tank-oil volume, with units scf/STB. \( R_s \) is constant at pressures greater than the bubblepoint and decreases as gas is liberated at pressures below the bubblepoint.

The producing GOR, \( R_p \), defines the instantaneous ratio of the total surface-gas volume produced divided by the total stock-tank-oil volume. At pressures greater than bubblepoint, \( R_p \) is constant and equal to \( R_s \) at bubblepoint. At pressures less than the bubblepoint, \( R_p \) may be equal to, less than, or greater than the \( R_s \) of the flowing reservoir oil. Typically, \( R_p \) will increase 10 to 20 times the initial \( R_s \) because of increasing gas mobility and decreasing oil mobility during pressure depletion.

The surface volume ratio for gas condensates is usually expressed as an oil/gas ratio (GOR), \( R_{sg} \),

\[
R_{sg} = \frac{V_{g,sc}}{V_o} = \frac{V_{g,sc}}{V_{o,sc}} \tag{3.19}
\]

The unit for \( R_{sg} \) is STB/scf or, more commonly, “barrels per million” (STB/MMscf). To avoid misinterpretation, it should be clearly specified whether the GOR includes natural gas liquids (NGL’s) in addition to stock-tank condensate. In most petroleum engineering calculations, NGL’s are not included in the GOR.

The ratio of surface oil to surface gas produced from a single-phase reservoir gas is referred to as the solution oil/gas ratio, \( R_{og} \). At pressures above the dewpoint, the producing GOR, \( R_{og} \), is constant and equal to \( R_{sg} \) at the dewpoint. At pressures below the dewpoint, \( R_{og} \) is typically equal to or just slightly greater than \( R_{og} \); the contribution of flowing reservoir oil to surface-oil production is negligible in most gas-condensate reservoirs.

In the definitions of \( R_p \) and \( R_{og} \), the total producing surface-gas volume equals the surface gas from the reservoir oil plus the solution gas from the reservoir gas; likewise, the total producing surface oil equals the stock-tank oil from the reservoir oil plus the condensate from the reservoir gas. Fig. 3.4 illustrates the behavior of \( R_p \), \( R_{og} \), and \( 1/R_{og} \) as a function of pressure.

### 3.2.6 Viscosity

Two types of viscosity are used in engineering calculations: dynamic viscosity, \( \mu \), and kinematic viscosity, \( \nu \). The definition of \( \mu \) for Newtonian flow (which most petroleum mixtures follow) is

\[
\mu = \frac{\tau}{\nu} \tag{3.20}
\]

where \( \tau \) is shear stress per unit area in the shear plane parallel to the direction of flow, \( \nu \) is velocity gradient perpendicular to the plane of shear, and \( \nu \) is units conversion from mass to force. The two viscosities are related by density, where \( \mu = \rho \nu \).

Most petroleum engineering applications use dynamic viscosity, which is the property reported in commercial laboratory studies. The unit of dynamic viscosity is centipoise (cp), or in SI units, mPa·s, where 1 cp = 1 mPa·s. Kinematic viscosity is usually reported in centistoke (cSt), which is obtained by dividing \( \mu \) in cp by \( \rho \) in g/cm³; the SI unit for \( \nu \) is mm²/s, which is numerically equivalent to centistoke. Fig. 3.5 shows oil, gas, and water viscosities for typical reservoir systems.

### 3.2.7 Diffusion Coefficients

In the absence of bulk flow, components in a single-phase mixture are transported according to gradients in concentration (i.e., chemical potential). Fick’s law for 1D molecular diffusion in a binary system is given by

\[
i_u = -D_{ij} \left( \frac{\partial C_i}{\partial x} \right) \tag{3.21}
\]

where \( i \) and \( u \) refer to the chemical species under consideration, \( D_{ij} \) is the binary diffusion coefficient, \( C_i \) is the molar concentration of Component \( i \), and \( x \) is distance.

Eq. 3.21 clearly shows that mass transfer by molecular diffusion can be significant for three reasons: (1) large diffusion coefficients, (2) large concentration differences, and (3) short distances. A combination of moderate diffusion coefficients, concentration gradients, and distance may also result in significant diffusive flow. Molecular diffusion is particularly important in naturally fractured reservoirs because of relatively short distances (e.g., small matrix block sizes).

Low-pressure binary diffusion coefficients for gases, \( D_{ij}^o \), are independent of composition and can be calculated accurately from fundamental gas theory (Chapman and Enskog), which are basically the same relations used to estimate low-pressure gas viscosity. No well-accepted method is available to correct \( D_{ij}^o \) for mixtures at high pressure, but two types of corresponding-states correlations have been proposed: \( D_{ij} = D_{ij}^o (\kappa T_p) \) and \( D_{ij} = D_{ij}^o (\kappa f) \).

At low pressures, diffusion coefficients are several orders of magnitude smaller in liquids than in gases. At reservoir conditions, the difference between gas and liquid diffusion coefficients may be less than one order of magnitude.

### 3.2.8 IFT

Interfacial forces act between equilibrium gas, oil, and water phases coexisting in the pores of a reservoir rock. These forces
are generally quantified in terms of IFT, \( \sigma \); units of \( \sigma \) are dynes/cm (or equivalently, \( \text{mN/m} \)). The magnitude of IFT varies from 0.1 to 0.3 dynes/cm for high-pressure gas/oil mixtures. Gas/oil capillary pressure, \( P_{\text{cp}} \), is usually considered proportional to IFT according to the Young-Laplace equation \( P_{\text{cp}} = 2\sigma / r \), where \( r \) is an average pore radius. Recovery mechanisms that are influenced by capillary pressure (e.g., gas injection in naturally fractured reservoirs) will necessarily be sensitive to IFT.

### 3.3 Gas Mixtures

This section gives correlations for PVT properties of natural gases, including the following:

1. Review of gas volumetric properties.
2. Z-factor correlations.
3. Gas pseudocritical properties.
4. Wellstream gravity of wet gases and gas condensates.
5. Gas viscosity.
6. Dewpoint pressure.
7. Total volume factor.

#### 3.3.1 Review of Gas Volumetric Properties

The properties of gas mixtures are well understood and have been accurately correlated for many years with graphical charts and EOS's based on extensive experimental data.16-19 The behavior of gases at low pressures was originally quantified on the basis of experimental work by Charles and Boyle, which resulted in the ideal-gas law,\(^3\)

\[
pV = nRT \tag{3.22}\]

where \( R \) is the universal gas constant given in Appendix A for various units (Table A-2). In customary units,

\[
R = 10.73146 \text{ psia} \times \frac{1}{R} \tag{3.23}\]

while for other units, \( R \) can be calculated from the relation

\[
R = 10.73146 \frac{P_{\text{stat}}}{\text{psia}} \left( \frac{R}{T_{\text{stat}}} \right) \frac{V_{\text{stat}}}{\text{ft}^3} \frac{\text{lbm}}{\text{mol}}. \tag{3.24}\]

For example, the gas constant for SPE-preferred SI units is given by

\[
R = 10.73146 \times \left( 6.894757 \text{ kPa} \times \frac{1}{R} \right) \times \left( 0.02831685 \text{ m}^3 \text{ kg}^{-1} \text{mol}^{-1} \right) = 8.3144 \text{ J} / \text{K} \text{ mol}^{-1} \tag{3.25}\]

The gas constant can also be expressed in terms of energy units (e.g., \( R = 8.3143 \text{ J} / \text{mol} \cdot \text{K} \)), note that \( J = \text{N} \cdot \text{m} = (\text{Nm}) \cdot \text{m} = \text{Pa} \cdot \text{m}^3 \). In this case, the conversion from one unit system to another is given by

\[
R = 8.3144 \frac{P_{\text{stat}}}{\text{psia}} \frac{T_{\text{stat}}}{\text{R}} \frac{V_{\text{stat}}}{\text{ft}^3} \frac{\text{lbm}}{\text{mol}}. \tag{3.26}\]

An ideal gas is a hypothetical mixture with molecules that are negligible in size and have no intermolecular forces. Real gases mimic the behavior of an ideal gas at low pressures and high temperatures because the mixture volume is much larger than the volume of the molecules making up the mixture. That is, the mean free path between molecules that are moving randomly within the total volume is very large and intermolecular forces are thus very small.

Most gases at low pressure follow the ideal-gas law. Application of the ideal-gas law results in two useful engineering approximations. First, the standard molar volume representing the volume occupied by one mole of gas at standard conditions is independent of the gas composition.

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\[
\left( \frac{V_{\text{ref}}}{V} \right) = \left( \frac{P_{\text{ref}}}{P} \right) = \left( \frac{R T_{\text{ref}}}{R T} \right) = \frac{10.73146(60 + 459.67)}{14.7} = 379.4 \text{ scf/ibm mol} = 23.69 \text{ m}^3 / \text{kmol}. \tag{3.27}\]

Second, the specific gravity of a gas directly reflects the gas molecular weight at standard conditions,

\[
y_g = \frac{\rho_{\text{gas}}}{\rho_{\text{air}}} = \frac{M_g}{M_{\text{air}}} = \frac{M_g}{28.97} \tag{3.28}\]

For gas mixtures at moderate to high pressure or at low temperature the ideal-gas law does not hold because the volume of the constituent molecules and their intermolecular forces strongly affect the volumetric behavior of the gas. Comparison of experimental data for real gases with the behavior predicted by the ideal-gas law shows significant deviations. The deviation from ideal behavior can be expressed as a factor, \( Z \), defined as the ratio of the actual volume of one mole of a real-gas mixture to the volume of one mole of an ideal gas,

\[
Z = \frac{V_{\text{gas}}}{V} \tag{3.29}\]

where \( Z \) is a dimensionless quantity. Terms used for \( Z \) include deviation factor, compressibility factor, and \( Z \) factor. Z factor is used in this monograph, as will the SPE reserve symbol \( Z \) (instead of the recommended SPE symbol \( z \)) to avoid confusion with the symbol \( z \) used for feed composition.

From Eqs. 3.32 and 3.29, we can write the real-gas law including the \( Z \) factor as

\[
pV = nZRT \tag{3.30}\]

which is the standard equation for describing the volumetric behavior of reservoir gases. Another form of the real-gas law written in terms of specific volume \((\zeta = 1/\rho)\) is

\[
\zeta V = ZRT/M \tag{3.31}\]

or, in terms of molar volume \((\nu = M/\rho)\),

\[
\nu v = ZRT \tag{3.32}\]

Z factor, defined by Eq. 3.30,

\[
Z = pV/nRT \tag{3.33}\]

is used for both phases in EOS applications (see Chap. 4). In this monograph we use both \( Z \) and \( Z \) for gases and \( Z\) for oils; \( Z \) without a subsript always implies the \( Z \) factor of a “gas-like” phase. All volumetric properties of gases can be derived from the real-gas law. Gas density is given by

\[
\rho = PM/\rho \tag{3.34}\]

or, in terms of gas specific gravity, by

\[
\rho_s = 28.97 \frac{P}{RT} \tag{3.35}\]

For wet-gas and gas-condensate mixtures, wellstream gravity, \( \gamma_w \), must be used instead of \( \gamma_g \) in Eq. 3.35. Gas density may range from 0.05 \( \text{lbm/ft}^3 \) at standard conditions to 30 \( \text{lbm/ft}^3 \) for high-pressure gases.

Gas molar volume, \( V_m \), is given by

\[
V_m = ZRT/p \tag{3.36}\]

where typical values of \( V_m \) at reservoir conditions range from 1 to 1.5 \( \text{ft}^3/\text{lbm mol} \) compared with 379 \( \text{ft}^3/\text{lbm mol} \) for gases at standard conditions. In Eqs. 3.30 through 3.36, \( R \) = universal gas constant.
Gas compressibility, \( c_g \), is given by

\[
\frac{c_g}{C_{0043}/C_{0042}} = \frac{1}{p} - \frac{1}{Z} \left( \frac{Z}{np} \right)_T
\]

(3.37)

For sweet natural gas (i.e., not containing H₂S) at pressures less than \( \approx 1,000 \) psia, the second term in Eq. 3.37 is negligible and \( c_g = 1/p \) is a reasonable approximation.

Gas volume factor, \( B_g \), is defined as the ratio of gas volume at specified \( p \) and \( T \) to the ideal-gas volume at standard conditions,

\[
B_g = \left( \frac{p_{sc}}{p} \right) \left( \frac{T_{sc}}{T} \right)
\]

(3.38)

For customary units (\( p_{sc} = 14.7 \) psia and \( T_{sc} = 520^\circ \text{R} \)), this is

\[
B_g = 0.02827 \left( \frac{T_{sc}}{T} \right)
\]

(3.39)

with temperature in \(^\circ\text{R}\) and pressure in psia. This definition of \( B_g \) assumes that the gas volume at \( p \) and \( T \) remains as a gas at standard conditions. For wet gases and gas condensates, the surface gas will not contain all the original gas mixture because liquid is produced after separation. For these mixtures, the traditional definition of \( B_g \) may still be useful; however, we refer to this quantity as a hypothetical wet-gas volume factor, \( B_{gw} \), which is calculated from Eq. 3.38.

Because \( B_g \) is inversely proportional to pressure, the inverse volume factor, \( b_g = 1/B_g \), is commonly used. For field units,

\[
b_g \text{ in scf/ft}^3 = 35.37 \frac{p}{T} \quad \text{and} \quad b_g \text{ in Mscf/bbl} = 0.1985 \frac{p}{T}
\]

(3.40a, 3.40b)

If the reservoir gas yields condensate at the surface, the dry-gas volume factor, \( B_{gd} \), is sometimes used.

\[
B_{gd} = \left( \frac{p_{sc}}{p} \right) \left( \frac{T_{sc}}{T} \right) \left( \frac{1}{F_g} \right)
\]

(3.41)

where \( F_g = \text{ratio of moles of surface gas, } n_g, \text{ to moles of wellstream mixture (i.e., reservoir gas, } n_r); \text{ see Eqs. 7.10 and 7.11 of Chap. 7.}

### 3.3.2 Z-Factor Correlations

Standing and Katz\(^4\) present a generalized Z-factor chart (Fig. 3.6), which has become an industry standard for predicting the volumetric behavior of natural gases. Many empirical equations and EOS’s have been fit to the original Standing-Katz chart. For example, Hall and Yarborough\(^21,22\) present an
accurate representation of the Standing-Katz chart using a Carnahan-Staingly hard-sphere EOS,
\[ Z = 4 \pi \rho_{cr} f(\rho/\rho_{cr}), \]  
(3.42)
where \( \rho = 0.06125 \times \exp[-1.2(1 - \rho)], \) where \( \tau = 1/T_{\text{cr}}. \)

The reduced-density parameter, \( \rho \) (the product of a van der Waals covolume and density), is obtained by solving
\[ f(\rho) = 0 = - ap_{cr} + \frac{y}{1 - y} \times \left( \frac{y + y^2 + y^3 - y^4}{1 - y} \right) \]
\[ - (14.76 - 9.76y + 4.58y^2) \]
\[ + (90.7 - 242.21 + 42.4 \times 1.18^{1.18 + 2.82}) \]  
(3.43)
with \( df(\rho)/dy \) equals to the definition of \( c_i \) given by
\[ \frac{df(\rho)}{dy} = \frac{a}{\rho_{cr}^2} \times \left( \frac{1}{\rho_{cr}} \times \frac{ap_{cr} \rho - y^2}{df(\rho)/dy} \right), \]
(3.45)
An initial value of \( \gamma = 0.001 \) can be used with a Newton-Raphson procedure, where convergence should be obtained in 3 to 10 iterations for \( f(\rho) = 1 \times 10^{-3}. \)

On the basis of Takacs's comparison of eight correlations representing the Standing-Katz chart, the Hall and Yarborough, the Dranich and Abou-Kassem equations give the most accurate representation for a broad range of temperatures and pressures. Both equations are valid for \( \gamma \leq 3 \) and \( 0.2 \leq p_i \leq 25 \) to 30.

For many petroleum engineering applications, the Brill and Beggs equation gives a satisfactory representation (\pm 1 to 2%) of the original Standing-Katz Z-factor chart for \( 1.2 < T < 2. \) Also, this equation can be solved explicitly for \( Z. \) The main limitations are that reduced temperature must be \( > 1.2 \) (\( \approx 80 \text{°F} \)) and reduced pressure should be \( < 15 \) (\( \approx 10,000 \text{psia} \)).

The Standing and Katz Z-factor correlation may require special treatment for wet gas and gas-condensate fluids containing significant amounts of heptanes-plus material and for gas mixtures with significant amounts of nonhydrocarbons. An apparent discrepancy in the Standing-Katz Z-factor chart for \( 1.05 < T < 1.15 \) has been “smoothed” in the Hall-Yarborough correlations. The Hall and Yarborough (or Dranich and Abou-Kassem) equation is recommended for most natural gases. With today's computing capabilities, choosing simple, less-reliable equations, such as the Brill and Beggs Z-factor, is normally unnecessary.

The Lee-Kesler, AGA 8.30 and DDMIX 30 correlations for Z factor were developed with multiconstant EOS's to give accurate volumetric predictions for both pure components and mixtures. They require more computation but are very accurate. These equations are particularly useful in custody-transfer calculations. They also are required for gases containing water and concentrations of nonhydrocarbons that exceed the limits of the Wichert and Katz method. \( 30,31 \)

3.3.3 Gas Pseudocritical Properties. Z factor, viscosity, and other gas properties have been correlated accurately with corresponding-states principles, where the property is correlated as a function of reduced pressure and temperature.
\[ \frac{Z}{T_{\text{cr}}} = f(p_{cr}, T_{\text{cr}}) \]
(3.46)
and \( \mu_{g}/\mu_{\text{hc}} = f(p_{cr}, T_{\text{cr}}). \)

where \( p_{cr} = p/p_c \) and \( T_c = T/T_c. \) Such corresponding-states relations should be valid for most pure compounds when component critical properties \( p_{cr} \) and \( T_c \) are used. The same relations can be used for gas mixtures if the mixture pseudocritical properties \( p_{cr} \) and \( T_c \) are used. Pseudocritical properties of gases can be estimated with gas composition and mixing rules or from correlations based on gas-specific gravity.

Sutton's following correlations for hydrocarbon gas mixtures.
\[ T_{pc} = 169.2 + 349.5 \gamma_{pc} - 74.0 \gamma_{pc}^2 \]  
(3.47a)
and \( p_{pc} = 756.8 - 131 \gamma_{pc} - 3.6 \gamma_{pc}^2 \)  
(3.47b)
He claims that Eqs. 3.47a and 3.47b are the most reliable correlations for calculating pseudocritical properties with the Standing-Katz Z-factor chart. He even claims that this method is superior to the use of composition and mixing rules.

Standing \( 3 \) gives two sets of correlations: one for dry hydrocarbon gases (\( \gamma_{pc} < 0.75 \)),
\[ T_{pc} = 168 + 325 \gamma_{pc} - 12.5 \gamma_{pc}^2 \]  
(3.48a)
and \( p_{pc} = 667 + 15.5 \gamma_{pc} - 37.5 \gamma_{pc}^2 \)  
(3.48b)
and one for wet-gas mixtures (\( \gamma_{pc} \geq 0.75 \)),
\[ T_{pc} = 187 + 330 \gamma_{pc} - 71.5 \gamma_{pc}^2 \]  
(3.49a)
and \( p_{pc} = 706 + 51.7 \gamma_{pc} - 11.3 \gamma_{pc}^2 \)  
(3.49b)

The Standing correlations are used extensively in the industry; Fig. 3.7 compares them with the Sutton correlations. The Sutton and the Standing wet-gas correlations for \( T_{pc} \) give basically the same results, whereas the three \( p_{pc} \) correlations are quite different at \( \gamma_{pc} > 0.85. \) Kay's mixing rule is typically used when gas composition is available.

\[ M = \sum_{i=1}^{N} \gamma_i M_i \]  
(3.50a)

---

**Fig. 3.7—Gas pseudocritical properties as functions of specific gravity.**

---

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and \( e = 120 \left( \frac{(y_{CO_2} + y_{H_2S})^{10} - (y_{CO_2} + y_{H_2S})^{4}}{4} \right) 
\left( \frac{y_{CO_2}^{4} - y_{H_2S}^{4}}{4} \right) \) \]

where \( T'_p \) and \( p'_p \) are mixture pseudocriticals based on Kay’s mixing rule. This method was developed from extensive data from natural gases containing nonhydrocarbons, with CO\(_2\) molar concentrations ranging from 0 to 55% and H\(_2\)S molar concentrations ranging from 0 to 74%. If only gas gravity and nonhydrocarbon content are known, the hydrocarbon specific gravity is first calculated from

\[
y_{HC} = \frac{y_{CO_2} + y_{H_2S} + y_{CO_2}M_{CO_2}C_7^* + y_{H_2S}M_{H_2S}C_7^*}}{M_{HC}}
\]

Hydrocarbon pseudocriticals are then calculated from Eqs. 3.47a and 3.47b, and these values are adjusted for nonhydrocarbon content on the basis of Kay’s mixing rule.

\[
p'_p = \left(1 - y_{CO_2} - y_{H_2S}\right) p_{ HC, p}
\]

and \( T'_p \) and \( p'_p \) are used in the Wichert-Aziz equations with CO\(_2\) and H\(_2\)S molar fractions to obtain mixture \( T_p \) and \( p_p \). The Sutton’s correlations (Eqs. 3.47a and 3.47b) are recommended for hydrocarbon pseudocritical properties. If composition is available, Kay’s mixing rule should be used with the Matthews et al. pseudocriticals for C\(_7\)+. Gases containing significant amounts of CO\(_2\) and H\(_2\)S nonhydrocarbons should always be corrected with the Wichert-Aziz equations. Finally, for gas-condensate fluids the wellstream specific gravity, \( \gamma_w \), (discussed in the next section), should replace \( \gamma_p \) in the equations above.

### 3.3.4 Wellstream Specific Gravity

Gas mixtures that produce condensate at surface conditions may exist as a single-phase gas in the reservoir and production tubing. This can be verified by determining the dewpoint pressure at the prevailing temperature. If wellstream properties are desired at conditions where the mixture is single-phase, surface-gas and -oil properties must be converted to a wellstream specific gravity, \( \gamma_w \). This gravity should be used instead of \( \gamma_p \) to estimate pseudocritical properties.

Wellstream gravity, \( \gamma_w \), represents the average molecular weight of the produced mixture (relative to air) and is readily calculated from the producing-oil (condensate)/gas ratio, \( r_{OG} \), average surface-gas gravity \( \gamma_p \); surface-condensate gravity, \( \gamma_o \); and surface-condensate molecular weight, \( M_p \).

\[
\gamma_w = \frac{T_p + 4,880 r_{OG} \gamma_o}{1 + 133,000 r_{OG} \gamma_o / M_p} \tag{3.55}
\]

with \( r_{OG} \) in STB/scf. Average surface-gas gravity is given by

\[
T_p = \frac{\Sigma \gamma_i T_{pi}}{\Sigma \gamma_i} \tag{3.56}
\]

where \( R_{pi} = GOR \) of Separator Stage \( i \). Standing\(^{31}\) presents Eq. 3.55 graphically in Fig. 3.9.

When \( M_p \) is not available, Standing gives the following correlation.
This relation should not be extrapolated outside the range 45 < \gamma_{AP} < 60. Eilerts\textsuperscript{34} gives a relation for \((\gamma/M)_{P}\),

\[
\frac{\gamma}{M} = \left[ 1.89 \times 10^{-3} \right] + \left( 7.35 \times 10^{-3} \right) \gamma_{AP} - \left( 4.52 \times 10^{-3} \right) \gamma_{AP}^2
\]

which should be reliable for most condensates. When condensate molecular weight is not available, the recommended correlation for \(M_{s}\) is the Cragoe\textsuperscript{35} correlation,

\[
M_{s} = 0.084 \frac{\gamma_{AP} - 5.9}{\gamma_{AP}}
\]

which gives reasonable values for all surface condensates and stock-tank oils.

A typical problem that often arises in the engineering of gas-condensate reservoirs is that all the data required to calculate wellstream gas volumes and wellstream specific gravity are not available and must be estimated.\textsuperscript{36-38} In practice, we often report only the first-stage-separator GOR (relative to stock-tank-oil volume) and gas specific gravity, \(R_{1}\) and \(\gamma_{1}\), respectively; the stock-tank-oil gravity, \(\gamma_{ST}\), and the primary-separator conditions, \(p_{w1}\) and \(T_{sp1}\).

To calculate \(\gamma_{1}\), from Eq. 3.55 we need total producing OGR, \(R_{p}\), which equals the inverse of \(R_{1}\) plus the additional gas that will be released from the first-stage separator oil, \(R_{1}\),

\[
r_{p} = \frac{1}{R_{1} + R_{1}}
\]

\(R_{1}\), can be estimated from several correlations.\textsuperscript{37,39} Whitson\textsuperscript{38} proposes use of a bubblepoint pressure correlation (e.g., the Standing\textsuperscript{40} correlation),

\[
R_{1} = A_{2} \gamma_{p} \hspace{1cm} \text{(3.61a)}
\]

and

\[
A_{1} = \left( \frac{p_{w1}}{18.2} + 1.4 \right) ^{10^{0.012} \gamma_{AP} - 0.0097 (\gamma_{AP})^{0.15}}
\]

(3.61b)

with \(p_{w1}\) in psia, \(T_{sp1}\) in \(^\circ\)F, and \(R_{1}\) in scf/STB. \(\gamma_{p}\) is the gas gravity of the additional solution gas released from the separator oil. The Katz\textsuperscript{41} correlation (Fig. 3.10) can be used to estimate \(\gamma_{p}\), where a best-fit representation of his graphical correlation is

\[
\gamma_{p} = A_{2} + A_{1} R_{1} \hspace{1cm} \text{(3.62)}
\]

where \(A_{2} = 0.25 + 0.02 \gamma_{AP}\) and \(A_{1} = -(3.57 \times 10^{-6} \gamma_{AP}^{2})\).

Solving Eqs. 3.61 and 3.62 for \(R_{1}\), gives

\[
R_{1} = \frac{A_{2} \gamma_{p} - A_{1}}{[1 - A_{1} \gamma_{p}]} \hspace{1cm} \text{(3.63)}
\]

Average surface separator gas gravity, \(\gamma_{S}\), is given by

\[
\gamma_{S} = \gamma_{p} R_{p} + \gamma_{1} \hspace{1cm} R_{1} = \frac{R_{p}}{R_{p} + R_{1}}
\]

Although the Katz correlation is only approximate, the impact of a few percent error in \(\gamma_{p}\) is not of practical consequence to the calculation of \(\gamma_{1}\) because \(R_{1}\) is usually much less than \(R_{p}\).

3.3.5 Gas Viscosity. Viscosity of reservoir gases generally ranges from 0.01 to 0.03 cp at standard and reservoir conditions, reaching up to 1 cp for near-critical gas condensates. Estimation of gas viscosities at elevated pressure and temperature is typically a two-step procedure: (1) calculating mixture low-pressure viscosity \(\mu_{low}\) at \(p_{w1}\) and \(T_{p}\) from Chapman-Enskog theory\textsuperscript{36,37} and (2) correcting this value for the effect of pressure and temperature with a corresponding-states or dense-gas correlation. These correlations relate the actual viscosity \(\mu_{G}\) at ps and \(T_{sp}\) to low-pressure viscosity by use of the ratio \(\mu_{G}/\mu_{low}\) or difference (\(\mu_{G} - \mu_{low}\)) as a function of pseudoreduced properties \(p_{r}\) and \(T_{r}\) or as a function of pseudoreduced density \(\rho_{r}\).

Gas viscosities are rarely measured because most laboratories do not have the required equipment; thus, the prediction of gas viscosity is particularly important. Gas viscosity of reservoir systems is often estimated from the graphical correlation \(\mu_{G}/\mu_{ew} = R_{low} p_{w}\) proposed by Carr et al.\textsuperscript{42} (Fig. 3.11). Dempsey\textsuperscript{43} gives a polynomial approximation of the Carr et al. correlation. With these correlations, gas viscosities can be estimated with an accuracy of about \(\pm 5\%\) for most applications. The Dempsey correlation is valid in the range 1.2 < \(T_{r}\) < 3 and 1 < \(p_{r}\) < 20.

The Lee-Gonzalez gas viscosity correlation (used by most PVT laboratories when reporting gas viscosities) is given by\textsuperscript{44}

\[
\mu_{G} = A_{1} \times 10^{-4} \exp\left( A_{2} p_{r}^{A_{3}} \right) \hspace{1cm} \text{(3.65a)}
\]

where

\[
A_{1} = \left[ \frac{9.379 + 0.01607 M_{s}}{10^{5.6}} \right] \hspace{1cm} \text{(3.65b)}
\]

\[
A_{2} = 3.448 + \frac{986.4}{T} + 0.01009 M_{s} \hspace{1cm} \text{(3.65b)}
\]

and

\[
A_{3} = 2.447 - 0.2224 A_{2} \hspace{1cm} \text{(3.65b)}
\]
Lucas proposes the following gas viscosity correlation, which is valid in the range $1/T_\text{r} < 40$ and $0 < p_\text{r} < 100$ (Fig. 3.12):\[{3.66}\]

$$
\mu/\mu_{sc} = 1 + A_1 (p_\text{r}/p_{pc})^3 \exp \left[ - \frac{5.1726}{T_\text{r}/T_{pc}} - 0.3286 \right],
$$

where

$$
A_1 = \frac{1.245 \times 10^{-3}}{T_{pc}^{0.3286}}.
$$

$$
A_2 = A_1 (1.6553 T_{pc} - 1.2723),
$$

$$
A_3 = 0.4489 \exp \left[ 3.057 T_{pc}^{0.77732} \right],
$$

$$
A_4 = 1.7368 \exp \left[ -0.18537 T_{pc}^{0.6351} \right],
$$

and $A_5 = 0.9425 \exp \left[ -0.18537 T_{pc}^{0.6351} \right].$

Given its wide range of applicability, the Lucas method is recommended for general use. When compositions are not available, correlations for pseudocritical properties in terms of specific gravity can be used instead. Standing gives equations for $\mu_{sc}$ in terms of $\gamma_s$, temperature, and nonhydrocarbon content,

$$
\mu_{sc} = (\mu_{sc})_{\text{assumed}} + \Delta \mu_{H_2S} + \Delta \mu_{CO_2} + \Delta \mu_{H_2O},
$$

where $\Delta \mu_{H_2S}$, $\Delta \mu_{CO_2}$, and $\Delta \mu_{H_2O}$ are corrections for $H_2S$, $CO_2$, and water, respectively.

**Fig. 3.11—Carr et al. gas-viscosity correlation.**

$$
\xi = 9.490 \left( \frac{T_{pc}}{M p_{pc}} \right)^{1/6},
$$

and $p_{pc} = RT_{pc}$

$$
\sum_{i=1}^{N} \gamma_i Z_i, \quad \sum_{i=1}^{N} V_i^\gamma, \quad \sum_{i=1}^{N} V_i^\gamma Z_i,
$$

with $\xi$ in cp, $T$ and $T_p$ in °R, and $p$ in psia. Special corrections should be applied to the Lucas correlation when polar compounds, such as $H_2S$ and water, are present in a gas mixture. The effect of $H_2S$ is always $<1\%$ and can be neglected, and appropriate corrections can be made to treat water if necessary.
where \( \mu_{\text{micro}} = (8.188 \times 10^{-1}) + (1.709 \times 10^{-1}) \)

\[-(2.062 \times 10^{-5})Z + (6.15 \times 10^{-5}) \log \gamma_p \]

\[\Delta \mu_{\text{CO}_2} = \gamma_{\text{CO}_2} \left[ (9.08 \times 10^{-1}) \log \gamma_p + (6.24 \times 10^{-1}) \right] \]

\[\Delta \mu_{\text{H}_2} = \gamma_{\text{H}_2} \left[ (8.49 \times 10^{-1}) \log \gamma_p + (3.73 \times 10^{-1}) \right] \]

Reid et al.\(^5\) review other gas viscosity correlations with accuracy similar to that of the Lucas correlation.

### 3.3.6 Dewpoint Pressure

The prediction of retrograde dewpoint pressure is not widely practiced. It is generally recognized that the complexity of retrograde phase behavior necessitates experimental determination of the dewpoint condition. Sage and Olds\(^{46}\) data are perhaps the most extensive tabular correlation of dewpoint pressures. Eilerts et al.\(^{47,48}\) also present dewpoint pressures for several light-condensate systems.

Nemeth and Kennedy\(^6\) have proposed a dewpoint correlation based on composition and \( C_{7s} \) properties.

\[
\ln p_d = A_1 \left[ z_{C_1} + z_{C_2} + 1.5 z_{C_3} + z_{C_4} + 2 z_{C_5} + z_{C_6} + z_{C_7} \right] + 0.4 z_{C_1} + 0.2 z_{C_2} + A_2 \gamma_{C_1} + A_3 \left[ \frac{z_{C_4}}{z_{C_4} + 0.002} \right] + A_4 T + A_5 z_{C_2} + A_6 z_{C_4} \]

\[+ A_7 \left[ \gamma_{C_4} + 0.0001 \right] + A_8 \left[ \gamma_{C_6} + 0.0001 \right] + A_9 \left[ \gamma_{C_8} + 0.0001 \right] \]

where \( A_1 = -2.0623054, A_2 = 6.6259728, A_3 = -4.4670559 \times 10^{-3}, A_4 = 1.4484364 \times 10^{-3}, A_5 = 3.267314 \times 10^{-2}, A_6 = -3.6453277 \times 10^{-3}, A_7 = 7.4299951 \times 10^{-5}, A_8 = -1.1381195 \times 10^{-1}, A_9 = 6.2476497 \times 10^{-3}, A_{10} = -1.0716866 \times 10^{-6}, \)

and \( A_{11} = 1.0746622 \times 10^{3} \).

The range of properties used to develop this correlation includes dewpoints from 1,000 to 10,000 psia, temperatures from 40 to 320°F, and a wide range of reservoir compositions. The correlation usually can be expected to predict dewpoints with an accuracy of ±10% for condensates that do not contain large amounts of nonhydrocarbons. This is acceptable in light of the fact that experimental dewpoint pressures are probably determined with an accuracy of only ±5%. The correlation is generally used only for preliminary reservoir studies conducted before an experimental dewpoint is available. Organick and Golding\(^{49}\) and Kurata and Kapl\(^{51}\) present graphical correlations for dewpoint pressure.

### 3.3.7 Total FVF

Total FVF,\(^{4,17,46}\) \( B_t \), is defined as the volume of a two-phase, gas-oil mixture (or sometimes a single-phase mixture) at elevated pressure and temperature divided by the stock-tank-oil volume resulting when the two-phase mixture is brought to surface conditions.

\[
B_t = \frac{F}{V_{ST}} = \frac{V_g + V_o}{V_{ST}} \quad \text{...................... (3.70)}
\]

\( B_t \) is used for calculating the oil in place for gas-condensate reservoirs, \( V_o \) is 0 in Eq. 3.70. Assuming 1 res bbl of hydrocarbon PV, the initial condensate in place is given by \( N = 1/B_t \) (in STB) and the initial "dry" separator gas in place is \( G = N/r_p \).

For gas-condensate systems, Sage and Olds\(^{46}\) give a tabulated correlation for \( B_t \).

\[
B_t = \frac{R_f T}{p R_g} Z' \quad \text{...................... (3.71)}
\]

where \( R_f \) is producing GOR in scf/STB, \( B_t \) is in bbl/STB, \( T \) is in °R, and \( p \) is in psia. \( Z' \) varies with pressure and temperature, where the tabulated correlation for \( Z' \) is well represented by

\[
Z' = A_0 + A_1 p_1 + A_2 p_2 + A_3 p_3 + A_4 p_4 + A_5 \frac{p^3}{T} \quad \text{...................... (3.72)}
\]

where \( A_0 = 5.050 \times 10^{-3}, A_1 = -2.740 \times 10^{-6}, A_2 = 3.331 \times 10^{-8}, A_3 = 2.198 \times 10^{-5}, A_4 = -2.675 \times 10^{-7} \) with \( p \) in psia and \( T \) in °R. Although the Sage and Olds data only cover the range 600 < \( p \) < 3,000 psia and 100 < \( T \) < 250°F, Eq. 3.72 gives acceptable results up to 10,000 psia and 350°F (when gas volume is much larger than oil volume).

When reservoir hydrocarbon volume consists only of gas, the following relations apply for total FVF:

\[
B_t = B_{p_{gas}} = B_{gas} / \left( R_o + C_m \right) \quad \text{...................... (3.73a)}
\]

\[
C_m = 133,000 / \left( \gamma_p / M_p \right) \quad \text{...................... (3.73b)}
\]

\[
M_p = 6.084 / (T_{ATR} - 5.9) \quad \text{...................... (3.73c)}
\]

\( \gamma_{p} \) is 141.5/131.5 + \( \gamma_p \) . \( \gamma_p \) is gas equivalent conversion factor in scf/STB (see Chap. 7), and \( R_o \) is producing GOR in scf/STB.

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Standing\textsuperscript{3} gives a graphical correlation for \( B_t \) using a correlation parameter \( A \) defined as
\[ A = R_{\text{Bo}}^{\frac{P_b T_b}{T}} , \]  
(3.74)
where \( a = 2.9 \times 10^{-0.00027 R_t} \). Standing’s correlation is valid for both oil and gas-condensate systems and can be represented with
\[ \log B_t = -5.262 - \frac{47.4}{12.22 + \log A^*}, \]  
(3.75a)
where \( \log A^* = \log A - (10.1 - \frac{96.8}{6.604 + \log p}) \) .  
(3.75b)
and \( A \) is given by Eq. 3.74. On the basis of data from North Sea oils, Glasø\textsuperscript{52} gives a correlation for \( B_t \) using the Standing correlation parameter \( A \) (Eq. 3.74):
\[ \log B_t = \left( 8.0135 \times 10^{-7} + 0.47257 \log A^* \right) + 0.17351 \left( \log A^* \right)^2, \]  
(3.76)
where \( A^* = A^{1.1089} \).

Either the Standing or the Glasø correlations for \( B_t \) can be used with approximately the same accuracy. However, neither correlation is consistent with the limiting conditions
\[ B_t = B_t^* \text{ for } V_g = 0, \]  
(3.77a)
and
\[ B_t = B_t^* R_p \text{ for } V_g = 0. \]  
(3.77b)

\( B_t \) correlations evaluated at a bubblepoint usually will underpredict the actual \( B_t \) by \( \approx 0.2 \).

### 3.4 Oil Mixtures

This section gives correlations for PVT properties of reservoir oils, including bubblepoint pressure and oil density, compressibility, FVF, and viscosity. With only a few exceptions, oil properties have been correlated in terms of surface-oil and -gas properties, including solution gas/oil ratio, oil gravity, average surface-gas gravity, and temperature. A few correlations are also given in terms of composition and component properties.

Reservoir oils typically contain dissolved gas consisting mainly of methane and ethane, some intermediates (C\textsubscript{3} through C\textsubscript{6}), and lesser quantities of nonhydrocarbons. The amount of dissolved gas has an important effect on oil properties. At the bubblepoint, a discontinuity in the system volumetric behavior is caused by gas coming out of solution, with the system compressibility changing dramatically.\textsuperscript{5} An accurate method is needed to correlate the bubblepoint pressure, temperature, and solution gas/oil ratio.

Oil properties can be grouped into two categories: saturated and undersaturated properties. Saturated properties apply at pressures at or below the bubblepoint, and undersaturated properties apply at pressures greater than the bubblepoint. For oils with initial GOR’s less than \( \approx 500 \text{ scf/STB} \), assuming linear variation of undersaturated-oil properties with pressure is usually acceptable.

#### 3.4.1 Bubblepoint Pressure

The correlation of bubblepoint pressure has received more attention than any other oil-property correlation. Standing\textsuperscript{1,17} developed the first accurate bubblepoint correlation, which was based on California crude oils.

The correlation of bubblepoint pressure, where mole fraction \( y_g \) of solution gas in the reservoir oil is used as the main correlating parameter\textsuperscript{12}:
\[ p_b = 18.2[A - 1.4], \]  
(3.78)
where \( A = \frac{R_t [y_g]^{0.57246}[0.000197 - 0.00259 y_g]}{[} \), with \( R_t \) in scf/STB, \( T \) in °F, and \( p_b \) in psia.

Lasater\textsuperscript{15} used a somewhat different approach to correlate bubblepoint pressure, where mole fraction \( y_g \) of solution gas in the reservoir oil is used as the main correlating parameter\textsuperscript{11}:
\[ p_b = A \frac{T}{y_g}, \]  
(3.79)
with \( T \) in °R and \( p_b \) in psia. The function \( A (y_g) \) is given graphically by Lasater, and his correlation can be accurately described by
\[ A = 0.83918 \times 10^{1.7764 y_g^0.227266}; \]  
(3.80a)
and
\[ A = 0.83918 \times 10^{1.0351 y_g^0.112104}; \]  
(3.80b)
where \( y_g = \left[ \frac{1 + 133.000 [y_g/M_o]}{R_t} \right]^2 \) .  
(3.81)

with \( R_t \) in scf/STB. In this correlation, the gas mole fraction is dependent mainly on solution gas/oil ratio, but also on the properties of the stock-tank oil. The Cracoge\textsuperscript{56} correlation given by Eq. 3.59 is recommended for estimating \( M_o \) when stock-tank-oil molecular weight is not known.

Standing’s approach was used by Glasø\textsuperscript{52} for North Sea oils, resulting in the correlation
\[ \log p_b = 1.7669 + 1.7447 \log A - 0.30218(\log A)^2, \]  
(3.82)
where \( A = \frac{R_t [y_g]^{0.57246}[0.000197 - 0.00259 y_g]}{[} \), with \( p_b \) in psia, \( T \) in °F, and \( R_t \) in scf/STB. Glasø’s corrections for nonhydrocarbon content and stock-tank-oil paraffinicity are not widely used, primarily be-
cause the necessary data are not available. Sutton and Farshad mention that the API correction for paraffinicity worsened bubblepoint predictions for gulf coast fluids. Fig. 3.13 gives an explanation for this observation.

Fig. 3.13 shows the effect of paraffinicity (which is quantified by the Watson characterization factor, \( K_w \)) on bubblepoint pressure; the figure is based on calculations with a tuned EOS for an Asian oil (solid circles). The oil composition is constant in the example calculation. The 12 \( C_{27} \) fractions are each split into a paraffinic pseudo-component and an aromatic pseudo-component (i.e., 24 \( C_{27} \) pseudo-components). The paraffinic fraction was varied, and bubblepoint calculations were made. The variation in paraffinicity is expressed in terms of the overall \( C_{27} \) Watson characterization factor. Also shown in the figure are the variation in solution gas/oil ratio and the oil specific gravity with \( K_w \). The actual reservoir oil has \( K_w = 11.55 \), where the EOS bubblepoint is close to the uncorrected Glasø bubblepoint prediction. When the correction for paraffinicity is applied, the correction gives a poorer bubblepoint prediction (even though the overall trend in bubblepoints is improved by the Glasø paraffinicity corrections).

A quantitatively similar correction to the Glasø correction (but easier to use) is based on the estimate for Whittington's Watson characterization factor, \( K_w \), and yields

\[
\gamma_{\text{API}}^m = \gamma_{\text{API}}^n \left[ \frac{1}{1 + 10^{-11.172 / \text{API}}} \right]^{0.1413}, \quad \text{(3.83)}
\]

The corrected specific gravity correction is used in the Glasø bubblepoint correlation instead of the measured specific gravity. An estimate of \( K_w \) for the stock-tank oil must be available to use this correction. Vazquez and Beggs give the following correlations. For \( \gamma_{\text{API}} \leq 30 \),

\[
p_g = \left[ 27.64 \left( \frac{100}{p_g} \right) \right]^{10^{-11.172 / \text{API}}} \left( T_p + 60 \right)^{0.1413}, \quad \text{(3.84)}
\]

and, for \( \gamma_{\text{API}} > 30 \),

\[
p_g = \left[ 56.06 \left( \frac{100}{p_g} \right) \right]^{10^{-11.172 / \text{API}}} \left( T_p + 60 \right)^{0.8425}, \quad \text{(3.85)}
\]

with \( p_g \) in psia, \( T_p \) in °F and \( R_i \) in scf/STB. These equations are based on a large number of data from commercial laboratories. Vazquez and Beggs correct for the effect of separator conditions using a modified gas specific gravity, \( \gamma_{ps} \), which is correlated with first-stage-separator pressure and temperature, and stock-tank-oil gravity.

\[
\gamma_{ps} = \gamma_{\text{API}} \left( 1 + 0.5912 \times 10^{-4} \gamma_{\text{API}} T_p \right) \left( \frac{p_{ps}}{14.7} \right), \quad \text{(3.86)}
\]

with \( T_p \) in °F and \( p_{ps} \) in psia. Standing's correlation can be used to develop field- or reservoir-specific bubblepoint correlations. A linear relation is usually assumed between bubblepoint pressure and the Standing correlating coefficient. This is a standard approach used in the industry, and the Standing bubblepoint correlating parameter is well suited for developing field-specific correlations. Sometimes the solution gas/oil ratio is needed at a given pressure, and this is readily calculated by solving the bubblepoint correlation for \( R_i \). For the Standing correlation,

\[
R_i = \gamma_{ps} \left( 0.0585 p_g + 1.41 \times 10^{10.096} / \text{API} \right) \left( \frac{T_p + 60}{14.7} \right)^{0.205}, \quad \text{(3.87)}
\]

similar relations can be derived for the other bubblepoint correlations.

In summary, significant differences in predicted bubblepoint pressures should not be expected for most reservoir oils with most of the previous correlations. The Lasater and Standing equations are recommended for general use and as a starting point for developing reservoir-specific correlations. Correlations developed for a specific region, such as Glasø's correlation for the North Sea, should probably be used in that region and, in the case of Glasø's correlation, may be extended to other regions by use of the paraffinicity correction.

### 3.4.2 Oil Density

Density of reservoir oil varies from 30 lbm/ft^3 for light volatile oils to 60 lbm/ft^3 for heavy crudes with little or no solution gas. Oil compressibility may range from \( 3 \times 10^{-6} \) psia for heavy crude oils to \( 50 \times 10^{-6} \) psia for light oils. The variation of oil compressibility with pressure is usually small, although for volatile oils the effect can be significant, particularly for material-balance and reservoir-simulation calculations of highly undersaturated volatile oils. Several methods have been used successfully to correlate oil volumetric properties, including extensions of ideal-solution mixing, EOS's, corresponding-states correlations, and empirical correlations. Oil density based on black-oil properties is given by

\[
\rho_o = \frac{62.4 \gamma_{ps} + 0.0136 \gamma_{ps} R_i}{B_o} \quad \text{(3.88)}
\]

with \( \rho_o \) in lbm/ft^3, \( B_o \) in bbl/STB, and \( R_i \) in scf/STB. Correlations can be used to estimate \( R_i \) and \( B_o \) from \( \gamma_{ps} \), \( \gamma_{API} \), \( T_p \), and \( p_{ps} \).

### Standing-Katz Method

Standing's3,17 and Standing and Katz give an accurate method for estimating oil densities that uses an extension of ideal-solution mixing.

\[
\rho_o = \rho_{ps} + \Delta \rho_o - \Delta \rho_{ps} \quad \text{(3.89)}
\]

where \( \rho_{ps} \) is the pseudoliquid density at standard conditions and the terms \( \Delta \rho_o \) and \( \Delta \rho_{ps} \) give corrections for temperature and pressure, respectively. Pseudoliquid density is calculated with ideal-solution mixing and correlations for the apparent liquid densities of ethane and methane.
GAS AND OIL PROPERTIES AND CORRELATIONS

Fig. 3.15—Chart for calculating pseudoliquid density of reservoir oil (from Standing 33).

and methane at standard conditions. Given oil composition \( x_i \), \( \rho_{po} \) is calculated from

\[
\rho_{po} = \frac{\sum_{i=1}^{n} x_i M_i}{\sum_{i=1}^{n} (x_i M_i / \rho_i)}, \quad \text{................................. (3.90)}
\]

where Standing and Katz show that apparent liquid densities \( \rho_i \) of \( C_2 \) and \( C_1 \) are functions of the densities \( \rho_{C_2} \) and \( \rho_{C_1} \), respectively (Fig. 3.14).

\[
\rho_{C_2} = 15.3 + 0.3167 \rho_{C_2+}, \quad \text{................................. (3.91)}
\]

\[
\rho_{C_1} = 0.312 + 0.45 \rho_{po}, \quad \text{................................. (3.92)}
\]

where \( \rho_{C_2+} = \sum_{i=C_2}^{C_7} x_i M_i / \rho_i \), ................................. (3.93)

with \( \rho \) in lbm/ft\(^3\). Application of these correlations results in an apparent trial-and-error calculation for \( \rho_{po} \). Standing 33 presents a graphical correlation (Fig. 3.15) based on these relations, where \( \rho_{po} \) is found from \( \rho_{C_2+} \) and weight fractions of \( C_2 \) and \( C_1 \) (\( w_{C_2} \) and \( w_{C_1} \), respectively).

Figs. 3.16 and 3.17 show the pressure and temperature corrections, \( \Delta \rho_p \) and \( \Delta \rho_T \), graphically. \( \Delta \rho_p \) is a function of \( \rho_{po} \), and \( \Delta \rho_T \) is a function of \( \rho_{po} / \rho_{C_2} \). Madrazo 59 introduced modified curves for \( \Delta \rho_p \) and \( \Delta \rho_T \) that improve predictions at higher pressures and temperatures. Standing 33 gives best-fit equations for his original graphical correlations of \( \Delta \rho_p \) and \( \Delta \rho_T \) (Eqs. 3.98 and 3.99), which are not recommended at temperatures > 240°F; instead, Madrazo’s graphical correlation can be used. The correction factors can also be used to determine isothermal compressibility and oil FVF at undersaturated conditions.

The treatment of nonhydrocarbons in the Standing-Katz method has not received much attention, and the method is not recommended when concentrations of nonhydrocarbons exceed 10 mol%. Standing 33 suggests that an apparent liquid density of 29.9 lbm/ft\(^3\) can be used for nitrogen but does not address how the nonhydrocarbons should be considered in the calculation procedure (i.e., as part of the C\(_{2+}\) material or following the calculation of \( \rho_{C_2} \) and \( \rho_{C_1} \)). Madrazo indicates that the volume contribution of nonhydrocarbons
Vogel and Yarborough\textsuperscript{60} suggest that the weight fraction of nitrogen should be added to the weight fraction of ethane. Using additive volumes and Eqs. 3.91 and 3.92, we can show that $\rho_{C_2}$ and $\rho_{po}$ can be calculated explicitly. Thus, the following is the most direct procedure for calculating $\rho_{po}$ from the Standing-Katz method.

1. Calculate the mass of each component.
   \[ m_i = x_i M_i \] \hspace{1cm} \text{(3.93)}

2. Calculate $V_{C_3}$.
   \[ V_{C_3} = \sum_{i=C_3}^{n} \frac{m_i}{\rho_i} \] \hspace{1cm} \text{(3.94)}
   where $\rho_i$ are component densities at standard conditions (Appendix A).

3. Calculate $\rho_{C_2}$.
   \[ \rho_{C_2} = -\frac{b + \sqrt{b^2 - 4ac}}{2a} \] \hspace{1cm} \text{(3.95)}

4. Calculate $V_{C_2}$.
   \[ V_{C_2} = V_{C_3} + \frac{m_{C_2}}{\rho_{C_2}} \]
   \[ = V_{C_3} + \frac{m_{C_2}}{15.3 + 0.3167\rho_{C_2}} \] \hspace{1cm} \text{(3.96)}

5. Calculate $\rho_{po}$.
   \[ \rho_{po} = \frac{-b + \sqrt{b^2 - 4ac}}{2a} \] \hspace{1cm} \text{(3.97)}
   where $a = 0.45V_{C_2}$, $b = m_{C_2} - 0.45m_{C_1} + 0.312V_{C_2}$, and $c = -0.312m_{C_2}$.

6. Calculate the pressure effect on density.
   \[ \Delta \rho_p = 10^{-7}[0.167 + (16.181 \times 10^{-0.0042p})] \rho \]
   \[ - 10^{-7}[0.299 + (263 \times 10^{-0.0042p})] \rho^2 \] \hspace{1cm} \text{(3.98)}

where $a = 0.3167V_{C_2}$, $b = m_{C_2} - 0.3167m_{C_2} + 15.3V_{C_2}$, and $c = -15.3m_{C_2}$.
7. Calculate the temperature effect on density.
\[
\Delta \rho_T = (T - 60) \left[ 0.0133 + 152.4(\rho_{po} + \Delta \rho_p)^{-2.45} - (T - 60) \left[ \left( \frac{152.4}{\rho_{po}} + \Delta \rho_p \right)^{2.45} - 8.1 (T - 60)^2 \right] \right] - 0.0622 \times 10^{-0.0136} \rho_{po} + \Delta \rho_p \right]. \tag{3.99}
\]

8. Calculate mixture density from Eq. 3.89.

In the absence of oil composition, Katz suggests calculating the pseudoliquid density from stock-tank-oil gravity, \( \rho_{po} \), solution gas/oil ratio, \( R_s \), and apparent liquid density of the surface gas, \( \rho_{ga} \), taken from a graphical correlation (Fig. 3.18).

\[
\rho_{po} = 38.52 \times 10^{-0.00126/\text{API}} + (94.75 - 33.93 \log \gamma_{API}) \log \gamma_e, \tag{3.101}
\]

with \( \rho_{po} \) in lbm/ft\(^3\) and \( R_s \) in scf/STB.

**Alani-Kennedy Method.** The Alani-Kennedy method for calculating oil density is a modification of the original van der Waals EOS, with constants \( a \) and \( b \) given as functions of temperature for normal paraffins \( C_1 \) to \( C_{10} \) and iso-butane (Table 3.1); two sets of coefficients are reported for methane (for temperatures from 70 to 300°F and from 301 to 460°F) and two sets for ethane (for temperatures from 100 to 249°F and from 250 to 460°F). Lohrenz et al.\(^{62}\) give Alani-Kennedy temperature-dependent coefficients for non-hydrocarbons \( N_2, CO_2, \) and \( H_2S \) (Table 3.1). The Alani-Kennedy equations are summarized next. Eqs. 3.102b and 3.102c are in the original van der Waals EOS but are not used.

\[
p = \frac{RT}{v - b} - a \frac{1}{v^2}, \tag{3.102a}
\]

**Standing gives an equation for** \( \rho_{po} \).
Fig. 3.18—Apparent pseudoliquid density of separator gas (from Standing,33 after Katz41).

\[ a_i = \frac{27}{64} \frac{RT^2}{P_c^2}, \quad (3.102b) \]
\[ b_i = \frac{1}{8} \frac{RT}{P_c}, \quad (3.102c) \]
\[ a = \sum_{i=1}^{\infty} x_i a_i, \quad (3.102d) \]
\[ b = \sum_{i=1}^{\infty} x_i b_i, \quad (3.102e) \]
\[ a_i = \frac{a_i}{T} + \log a_i; \quad i \neq C_7, \quad (3.102f) \]
\[ b_i = b_i T + b_{i2}; \quad i \neq C_7, \quad (3.102g) \]

where \( \log a_{C_7} = \left(3.8405985 \times 10^{-7}\right) M_{C_7} \)
\[-9.5638281 \times 10^{-6} M_{C_7} + \frac{261.80818}{T} + \left(7.310464 \times 10^{-5}\right) M_{C_7} + 10.753517 \]
\[ b_{C_7} = [3.499274 \times 10^{-7} M_{C_7} - 7.275403 \frac{1}{T}] + \left(2.232395 \times 10^{-7}\right) T - \left(1.632257 \times 10^{-7}\right) \frac{M_{C_7}}{T} + 6.2256545, \quad (3.103b) \]

with \( \rho \) in lbm/ft³, \( v \) in ft³/lbm mol, \( T \) in °R, \( p \) in psia, and \( R \) = universal gas constant = 10.73.

Solution of the cubic equation for volume is presented in Chap. 4. Density is given by \( \rho = \frac{M}{v} \), where \( M \) is the mixture molecular weight and \( v \) is the molar volume given by the solution to the cubic equation. The Alani-Kennedy method can also be used to estimate oil compressibilities.

Rackett,63 Hankinson and Thomson,64 and Hankinson et al.65 give accurate correlations for pure-component saturated-liquid densities, and although these correlations can be extended to mixtures, they have not been tested extensively for reservoir systems. Cullick et al.66 give a modified corresponding-states method for predicting density of reservoir fluids. The method has a better foundation and extrapolating capability than the methods discussed previously (particularly for systems with nonhydrocarbons); however, space does not allow presentation of the method in its entirety.

Either the Standing-Katz or Alani-Kennedy method should estimate the densities of most reservoir oils with an accuracy of ±2%. The Alani-Kennedy method is suggested for systems at temperatures >250°F and for systems containing appreciable amounts of nonhydrocarbons (>5 mol%). Cubic EOS’s (e.g., Peng-Robinson or Soave-Redlich-Kwong) that use volume translation also estimate liquid densities with an accuracy of a few percent (e.g., the recommended characterization procedures in Chap. 5 or other proposed characterizations67,68).

### Table 3.1—Constants for Alani-Kennedy61 Oil Density Correlation

<table>
<thead>
<tr>
<th>Component</th>
<th>( a_1 )</th>
<th>( a_2 )</th>
<th>( b_1 \times 10^5 )</th>
<th>( b_2 )</th>
</tr>
</thead>
<tbody>
<tr>
<td>N₂</td>
<td>4.300</td>
<td>2.293</td>
<td>4.49</td>
<td>0.3853</td>
</tr>
<tr>
<td>CO₂</td>
<td>8.166</td>
<td>126.0</td>
<td>0.1818</td>
<td>0.3872</td>
</tr>
<tr>
<td>H₂S</td>
<td>13.200</td>
<td>0.0</td>
<td>17.9</td>
<td>0.3945</td>
</tr>
<tr>
<td>C₁</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>At 70 to 300°F</td>
<td>9.1606413</td>
<td>61.893223</td>
<td>-3.3162472</td>
<td>0.50874303</td>
</tr>
<tr>
<td>At 300 to 460°F</td>
<td>147.47335</td>
<td>3.2474533</td>
<td>-14.072637</td>
<td>1.83286965</td>
</tr>
<tr>
<td>C₂</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>At 100 to 250°F</td>
<td>46.709573</td>
<td>-404.484844</td>
<td>5.1520981</td>
<td>0.52239654</td>
</tr>
<tr>
<td>At 250 to 460°F</td>
<td>17.495343</td>
<td>34.163551</td>
<td>2.8201736</td>
<td>0.62309877</td>
</tr>
<tr>
<td>C₃</td>
<td>20.24775</td>
<td>190.24420</td>
<td>2.1586448</td>
<td>0.90832519</td>
</tr>
<tr>
<td>H₂C₄</td>
<td>32.294420</td>
<td>131.63171</td>
<td>3.3862288</td>
<td>1.1013834</td>
</tr>
<tr>
<td>n-C₄</td>
<td>33.016212</td>
<td>146.15444</td>
<td>2.9021578</td>
<td>1.168144</td>
</tr>
<tr>
<td>i-C₄</td>
<td>37.046234</td>
<td>299.62630</td>
<td>2.1954785</td>
<td>1.4364289</td>
</tr>
<tr>
<td>n-C₅</td>
<td>37.046234</td>
<td>299.62630</td>
<td>2.1954785</td>
<td>1.4364289</td>
</tr>
<tr>
<td>i-C₅</td>
<td>52.093006</td>
<td>254.56097</td>
<td>3.6961858</td>
<td>1.5929406</td>
</tr>
<tr>
<td>n-C₆</td>
<td>82.296578</td>
<td>64.380912</td>
<td>5.2577968</td>
<td>1.7299902</td>
</tr>
<tr>
<td>i-C₆</td>
<td>89.185432</td>
<td>149.39026</td>
<td>5.9697530</td>
<td>1.9310993</td>
</tr>
<tr>
<td>n-C₇</td>
<td>124.062650</td>
<td>37.917238</td>
<td>6.7299934</td>
<td>2.1519973</td>
</tr>
<tr>
<td>i-C₇</td>
<td>146.643830</td>
<td>26.524103</td>
<td>7.8561789</td>
<td>2.3329874</td>
</tr>
</tbody>
</table>

PHASE BEHAVIOR
3.4.3 Undersaturated-Oil Compressibility. With measured data or an appropriate correlation for \( R_o \) or \( \rho_o \), Eq. 3.14 readily defines the isothermal compressibility of an oil at pressures greater than the bubblepoint. “Instantaneous” undersaturated-oil compressibility, defined by Eq. 3.15 with the pressure derivative evaluated at a specific pressure, is used in reservoir simulation and well-test interpretation. Another definition of oil compressibility may be used in material-balance calculations (e.g., Craft and Hawkins69)—the “cumulative” or “average” compressibility defines the cumulative volumetric change of oil from the initial reservoir pressure to current reservoir pressure.

\[
\sigma_c(p) = \frac{\int_c \sigma_c(p) \, dp}{V_o(\rho_o - p)} = -\frac{1}{V_o} \left[ \frac{V_o(p) - V_o}{\rho - \rho_o} \right].
\]  

(3.104)

The cumulative compressibility is readily identified because it is multiplied by the cumulative reservoir pressure drop, \( \rho - \rho_o \). Usually, \( \sigma_c \) is assumed constant; however, this assumption may not be justified for high-pressure volatile oils. Oil compressibility is used to calculate the variation in undersaturated density and FVF with pressure.

\[
\rho_o = \rho_o \exp[c_o(p - \rho_o)]
\]

\[
= \rho_o [1 - c_o(p - \rho_o)]
\]  

(3.105a)

where consistent units must be used. These equations are derived from the definition of isothermal compressibility assuming that \( c_o \) is constant. When oil compressibility varies significantly with pressure, Eqs. 3.105a and 3.105b are not really valid. The approximations \( \rho_o = \rho_o [1 - c_d(p - \rho_o)] \) and \( \rho_o = \rho_o [1 - c_d(p - \rho_o)] \) are used in many applications, and to predict volumetric behavior correctly with these relations requires that \( c_o \) be defined by

\[
c_o(p) = -\frac{1}{V_o} \left[ \frac{V_o(p) - V_o}{\rho - \rho_o} \right].
\]  

(3.106)

Strictly speaking, the compressibility of an oil mixture is defined only for pressures greater than the bubblepoint pressure. If an oil is at its bubblepoint, the compressibility can be determined and defined only for a positive change in pressure. A reduction in pressure from the bubblepoint results in gas coming out of solution and, subsequently, a change in the mass of the original system for which compressibility is to be determined. Implicit in the definition of compressibility is that the system mass remains constant. Vázquez and Beggs77 propose the following correlation for instantaneous undersaturated-oil compressibility.

\[
c_o = A/p. \]  

(3.107)

where

\[
A = 10^{-1}(9.8 + 17.2T - 1.180\rho_o + 12.61\gamma_{o,1} - 1.433),
\]

with \( c_o \) in psi \(^{-1} \), \( R_o \) in scf/STB, \( T \) in °F, and \( p \) in psia. With this correlation for oil compressibility, undersaturated-oil FVF can be calculated analytically from

\[
B_o = B_o(p_o/p_o)^4. \]  

(3.108)

If measured pressure/volume data are available (see Sec. 6.4 in Chap. 6), these data can be used to determine \( A \) (e.g., by plotting \( V_o/V_o \) vs. \( p/p_o \) on log-log paper). Constant \( A \) can then be used to compute compressibilities from the simple relation \( c_o = A/p \).

![Fig. 3.19—Undersaturated-oil-compressibility correlation (from Standing79).](image)

Constant \( A \) determined in this way is a useful correlating parameter, one that helps to identify erroneous undersaturated \( p-V_o \) data. Standing79 gives a graphical correlation for undersaturated \( c_o \) (Fig. 3.19) that can be represented by

\[
c_o = 10^{-6} \exp \left[ \frac{\rho_o + 0.004347(p - \rho_o) - 79.1}{(7.141 \times 10^{-6})(p - \rho_o) - 12.938} \right].
\]  

(3.109)

with \( c_o \) in psi \(^{-1} \), \( \rho_o \) in lbm/ft\(^3 \), and \( p \) in psia. The Alani-Kennedy EOS can also be solved analytically for oil compressibility, and Trube79 gives a corresponding-states method for determining oil compressibility with charts.

Any of the correlations mentioned here should yield reasonable estimates of \( c_o \). However, we recommend that experimental data be used for volatile oils when \( c_o \) is greater than about 20 \( \times 10^{-6} \) psi \(^{-1} \). A simple polynomial fit of the relative volume data, \( V_o/V_o \), from a PVT report allows an accurate and explicit equation for undersaturated-oil compressibility.

\[
V_o = A_o + A_1p + A_2p^2. \]  

(3.110a)

and

\[
c_o = -\frac{1}{V_o} \left( \frac{\partial V_o}{\partial p} \right)_p
\]

\[
= -\left( A_o + 2A_1p \right)\]  

(3.110b)

where \( A_o, A_1, \) and \( A_2 \) are determined from experimental data. Alternatively, measured data can be fit by use of Eq. 3.108.

3.4.4 Bubblepoint-Oil FVF. Oil FVF ranges from 1 bbl/STB for oils containing little solution gas to about 2.5 bbl/STB for volatile oils. \( B_o \) increases more or less linearly with the amount of gas in solution, a fact which explains why \( B_o \) correlates so well with bubblepoint pressure correlations. For example, Standing’s3,17,40 correlation for California crude oils is

\[
B_o = 0.9759 + (12 \times 10^{-5})A^{1.2}. \]  

(3.111)

where \( A = R_T[p_f/\gamma_o]^{0.5} + 1.25T \).
Glasp’s correlation for North Sea crude oils is
\[
\log(B_{ao} - 1) = -6.585 + 2.9133 \log A - 0.2768(\log A)^2,
\]
(3.112)
where
\[
A = R_{o} / (\gamma_1 / \gamma_2)^{0.829} + 0.9687 T.
\]
The Vazquez and Beggs\textsuperscript{57} correlation, based on data from commercial laboratories, is
\[
B_{ao} = 1 + (4.677 \times 10^{-6}) R_{o} + (0.1751 \times 10^{-3})(T - 60) \gamma_{ao}^{1.33} / \gamma_{st}^{0.75}
\]
(3.113a)
for \(\gamma_{st} \leq 30\) and
\[
B_{ao} = 1 + (4.67 \times 10^{-6}) R_{o} + (0.11 \times 10^{-3})(T - 60) \gamma_{ao}^{1.33} / \gamma_{st}^{0.75}
\]
(3.113b)
for \(\gamma_{st} > 30\), where the effect of separator conditions is included by use of a corrected gas gravity \(\gamma_{ao}\) (Eq. 3.86).

The Standing and the Vazquez-Beggs correlations indicate that a plot of \(R_{o}\) vs. \(R_{st}\) should correlate almost linearly. This plot is useful for checking the consistency of reported PVT data from a differential liberation plot. Eq. 3.114,\textsuperscript{1} which performs considerably better for Middle Eastern oils, also suggests a linear relationship between
\[
\frac{1}{R_{o}} = \frac{1}{R_{st}} + \frac{1}{R_{ao}}.
\]

3.4.5 Saturated-Oil Compressibility. Perrine\textsuperscript{6} introduces a definition for the compressibility of a saturated oil that includes the shrinkage effect of saturated-oil PVT, \(dR_{o}/dp\), and the expansion effect of gas coming out of solution, \(B_{o} R_{o}/p_R\).
\[
c_r = -\frac{1}{R_{o}} \frac{dR_{o}}{dp} + \frac{1}{3.515} \frac{B_{o} R_{o}}{p_R},
\]
(3.115)
where \(c_r\) is used in the definition of total system compressibility, \(c_s\).
\[
c_s = c_r S_o + c_o S_o + c_s S_t,
\]
(3.116)
where \(c_j\) relates to rock compressibility, \(c_r\) is in scf/STB, and \(c_s\) in bbl/STB = saturated-oil PVT at the pressure of interest, or below the original oil’s bubblepoint pressure (where both gas and oil are present).

3.4.6 Oil Viscosity. Typical oil viscosities range from 0.1 cp for near-critical oils to > 100 cp for heavy crudes. Temperature, stock-tank-oil density, and dissolved gas are the key parameters determining oil viscosity, where viscosity decreases with decreasing stock-tank-oil density (increasing oil temperature), increasing temperature, and increasing solution gas.

Oil viscosity is one of the most difficult properties to estimate, and most methods offer an accuracy of only about 10 to 20%. Two approaches are used to estimate oil viscosity: empirical and corresponding-states correlations. The empirical methods correlate gas-saturated oil viscosity in terms of dead-oil (residual or stock-tank-oil) viscosity and solution gas/oil ratio. Undersaturated oil viscosity is related to bubblepoint viscosity and the ratio or difference in actual and bubblepoint pressures. Corresponding-states methods use reduced density or reduced pressure and temperature as correlating parameters.

3.4.7 Dead-Oil (Residual- or Stock-Tank-Oil) Viscosity. Several correlations for dead-oil viscosity are given in terms of oil gravity and temperature. Standing,\textsuperscript{3} for example, gives best-fit equations for the original Beal\textsuperscript{72} graphical correlation.
\[
\mu_{ao} = \left(0.32 + 1.8 \times 10^{0.133 \gamma_{ao}^{0.33}}\right) \frac{360}{T + 200} \gamma_{ao},
\]
where \(A = 10^{0.813(6.824 - 0.04685 \gamma_{ao})}\).

A somewhat modified version of the original correlation is given in Fig. 3.20 by Standing,\textsuperscript{3} Beggs and Robinson,\textsuperscript{1} who give the following equation for the original Beal correlation,
\[
\mu_{ao} = -1 + 10^{-0.1384 \gamma_{ao}^{0.33} \gamma_{ao}}
\]
(3.118)
Bergman\textsuperscript{9} claims that the temperature dependence of the Beggs and Robinson correlation is not valid at lower temperatures (< 70°F) and suggests the following correlation, based on viscosity data, for pure compounds and reservoir oils.
\[
\log(A) = 3.141 \times 10^{10}(T - 1010) (\log \gamma_{ao}^{10.31})^{0.142} (\log T - 36.447),
\]
(3.120)
Al-Khafaji et al.\textsuperscript{74} give the following correlation
\[
\mu_{ao} = \frac{T_{10}^{0.0353 - 0.0048 T} \gamma_{ao}^{1.66}}{\gamma_{ao}^{0.33} + (T/30 - 14.29)^{1.78}},
\]
(3.121)
with \(T\) in °F and \(\mu_{ao}\) in cp for Eqs. 3.117 through 3.121.

Standing75 gives a relation for dead-oil viscosity in terms of dead-oil density, temperature, and the Watson characterization factor.

$$\log(\mu_{\text{do}} / \rho_o) = \frac{1}{A_1} T^2 \left[ K_o - \left( \frac{8.24}{\gamma_o} \right) + 1.639 A_2 - 1.059 \right] - 2.17,$$

where

- $$A_1 = 1 + 8.69 \log \frac{T + 460}{560}$$
- $$A_2 = 1 + 0.554 \log \frac{T + 460}{560}$$
- $$A_3 = -0.1285 \frac{2.87 A_2 - 1}{2.87 A_2 - \gamma_o}$$

and $$\mu_o = \frac{V_o}{1 + 0.000321(T - 60)10^{-0.00812 \text{API}}}.$$  

with $$\mu$$ in cp, $$T$$ in °F, and $$\rho$$ in g/cm³ for Eqs. 3.117 through 3.122. Eqs. 3.122a through 3.122e represent a best fit of the nomograph for viscosity in terms of temperature, gravity, and characterization factor. Eq. 3.122e (at standard pressure and temperature) is a best fit of thermal expansion data for crude oils.

Dead-oil viscosity is one of the most unreliable properties to predict with correlations primarily because of the large effect that oil type (paraffinicity, aromaticity, and asphaltene content) has on viscosity. For example, the oil viscosity of a crude oil with $$K_o = 12$$ may be 3 to 100 times the viscosity of a less paraffinic crude oil having the same gravity and $$K_o = 11$$. For this reason the Standing correlation based on the Watson characterization factor is recommended when $$K_o$$ is known. Using an incorrectly estimated $$K_o$$, however, may lead to a potentially large error in dead-oil viscosity.

### 3.4.8 Bubblepoint-Oil Viscosity

The original approach by Chew and Connally76 for correlating saturated-oil viscosity in terms of dead-oil viscosity and solution gas/oil ratio is still widely used.

$$\mu_{\text{so}} = A_1 (\mu_{\text{do}})^{A_2}.$$  

Fig. 3.22 shows the variation in $$\mu_{\text{so}}$$ with $$\mu_{\text{do}}$$ as a function of $$R_s$$.

The functional relations for $$A_1$$ and $$A_2$$ reported by various authors differ somewhat, but most are best-fit equations of Chew and Connally’s tabulated results.

**Beggs and Robinson.**

$$A_1 = 10.715(R_s + 100)^{0.515}$$

and $$A_2 = 5.44(R_s + 150)^{-0.308}.$$  

**Bergman.**

$$A_1 = 4.768 + 0.8359 \ln(R_s + 300)$$

and $$A_2 = 0.555 + 133.5 \frac{R_s}{R_s + 300}.$$  

**Standing.**

$$A_1 = 10^{-[7.4 \times 10^{-9} R_s + (2.2 \times 10^{-7} R_s)]}$$

and $$A_2 = \frac{0.68}{10^{0.62 \times 10^{-7} R_s}} + \frac{0.25}{10^{0.1 \times 10^{-7} R_s}} + \frac{0.062}{10^{0.74 \times 10^{-7} R_s}}.$$  

**Aziz et al.**

$$A_1 = 0.20 + (0.80 \times 10^{-0.00081 R_s})$$

and $$A_2 = 0.43 + (0.57 \times 10^{-0.00072 R_s}).$$

*Personal communication with D.F. Bergman, Amoco Research, Tulsa, Oklahoma (1992).*
Al-Khafaji et al.7 extend the Chew-Connally76 correlation to higher
gor’s (up to 2,000 scf/STB):

\[ A_1 = 0.247 + 0.282 A_0 + 0.5657 A_0^2 \]

\[- 0.4065 A_0^3 + 0.0631 A_0^4 \]  

(3.128a)  

and \[ A_2 = 0.894 + 0.0546 A_0 + 0.0766 A_0^2 \]

\[- 0.0736 A_0^3 + 0.01008 A_0^4 \]  

(3.128b)  

where \( A_1 \) = log(\( R_g \)) and \( R_g = 0.1 \) yields \( A_1 = A_0 = 1. R_g \) is
given in scf/STB for Eqs. 3.124 through 3.128. Chew and Connally indicate
that their correlation is based primarily on data with gor’s of < 1,000 scf/STB
and that the scatter in \( A_1 \) at higher gor’s is probably the result of insufficient
data. Eqs. 3.128a and 3.128b are based on additional data at higher gor’s. Eqs. 3.127a and 3.128b appear
to be the most well behaved.

An interesting observation by Abu-Khamsin and Al-Mahroon78 is
that saturated-oil viscosity, \( \mu_{so} \), correlates very well with satu-
rated-oil density, \( \rho_{so} \):

\[ \ln \rho_{so} = -2.652294 + 8.484462 \rho_{so}^{0.4} \]  

(3.129)  

with \( \rho_{so} \) in g/cm³. This behavior is expected from the Lohrenz et al.82 correlation discussed later. Although Abu-Khamsin and Al-
Mahroon do not comment on the applicability of Eq. 3.129 to under-
saturated oils, it would seem reasonable that their correlation should
apply to both saturated and undersaturated oils. In fact, the correla-
tion even appears to predict accurately dead-oil viscosities, \( \mu_{do} \), ex-
cept at low temperatures. Simon and Graue give graphical correlations
for the viscosity of saturated CO₂/oil systems (see Chap. 8).79

3.4.9 Undersaturated-Oil Viscosities. Beal72 gives the variation of
undersaturated-oil viscosity with pressure graphically where it has
been curve fit by Standing.5

\[ \mu_u = \mu_{so} (1 - (\rho - \rho_{so}) / \rho_{so})^{0.96} \]  

(3.130)  

The Vazquez and Beggs57 correlation is

\[ \mu_u = \mu_{so} \left( 1 + 0.0092545 \ln (\rho_{so}) \right) \]  

(3.131)  

where \( A = 2.6 \times 10^{-7} \exp \left[ -11.513 - (8.98 \times 10^{-3}) p_i \right] \).

A more recent correlation by Abdul-Majeed et al.80 is

\[ \mu_u = \mu_{so} + 0.1 (1 - 5.2106 + 1.11 \ln \rho_{so}) \]  

(3.132a)  

where \( A = 1.9311 - 0.89941 \ln \rho_{so} - 0.001194 \rho_{so}^{0.36} \)

+ 0.0092545 \ln (\rho_{so})  

(3.132b)  

Eq. 3.132 is based on the observation that a plot of \( \log (\mu_u - \mu_{so}) \)
vs. \( \log (\rho - \rho_{so}) \) plots as a straight line with slope of \( \approx 1.11 \). Be-
cause this observation appears to be fairly general, it can be used
to check reported undersaturated-oil viscosities and to develop
field-specific correlations.

Sutton and Farhadi84 and Khan et al.81 present results that indicate that
the Standing equation gives good results and that the Vaz-
quz-Beggs correlation tends to overpredict viscosities somewhat.
Abdul-Majeed et al.80 indicate that both the Standing and Vaz-
quz-Beggs correlations overpredict viscosities of North African and
Middle Eastern Oils (253 data), and that their own correlation per-
formed best for these data and for the data used by Vazquez
and Beggs.

3.4.10 Compositional Correlation. In compositional reservoir
simulation of miscible-gas-injection processes and the depletion of
near-critical reservoir fluids, the oil and gas compositions may be
very similar. A single viscosity relation consistent for both phases

is therefore desired. Several corresponding-states viscosity correla-
tions can be used for both oil and high-pressure gas mixtures; the
Lohrenz et al.82 correlation has become a standard in compositional
reservoir simulation. Lohrenz et al., the Jossi et al.82 correlation for
dense-gas mixtures (\( \rho_g > 0.1 \)),

\[ \left[ \mu - \mu \right] p_i^{0.81} = 0.10230 \pm 0.023364 \rho_{so} \]

+ 0.05853 \rho_{so} - 0.40475 \rho_{so} \]

+ 0.009324 \rho_{so} \]  

(3.133a)  

where \( \xi = 5.35 \ln \left( T_{pc} / M_p r_i \right) \)  

(3.133b)  

\( \rho_{so} = \rho_{so} / M_r \)  

(3.133c)  

and \( \rho_{so} = \rho_{so} / M_r \)  

(3.133d)  

Pseudocritical properties \( T_{pc}, P_{pc} \), and \( \rho_{so} \) are calculated
with Kay’s mixing rule.

Component viscosities, \( \mu_i \), can be calculated from the Lucas45
low-pressure correlation Eq. 3.67 or from the Stiel and Thodos83
 correlation (as suggested by Lohrenz et al.82).

\[ \mu_i = \left( 34 \times 10^{-7} \right) T_{hi}^{1.14} \]  

(3.134a)  

for \( T_{hi} \leq 1.5, \)

\[ \mu_i = \left( 17.78 \times 10^{-7} \right) \left( 4.58 T_{hi} - 1.67 \right)^{1/6} \]  

(3.134b)  

for \( T_{hi} > 1.5, \) where \( \xi = 5.35 \ln \left( T_{pc} / P_{pc} \right) \)  

(3.135b)  

Lohrenz et al.82 give a special relation for \( \rho_{so} \) of the C₂₂ fraction.

\[ \rho_{so} = 21.573 + 0.015122 \rho_{so} - 27.656 \rho_{so} + 0.070615 \rho_{so} \]  

(3.135c)  

with \( \rho \) in cp, \( \xi \) in cp⁻¹, \( \phi \) in lbm/lbm mol, \( T_{hi} \) in R, \( T \) in psia, and \( M_r \) in lbm/lbm mol. The Lohrenz et al. method is very
sensitive to mixture density and to the critical volumes of heavy
components. Adjustment of the critical volumes of heavy (and
sometimes light) components to match experimental oil viscosi-
ties is usually necessary.

3.5 IFT and Diffusion Coefficients

3.5.1 IFT. Weinaug and Katz84 propose an extension of the Ma-
cleod85 relationship for multicomponent mixtures.

\[ \sigma_{ij} = \sum_{i=1}^{N} \left( P_i \rho_i \right) \]  

(3.136)  

with \( \sigma \) in dynes/cm (mN/m) and \( \rho \) in g/cm³. \( P_i \) is the paracor of
Component \( i \), which can be calculated by group contributions, as
shown in Table 3.2. For n-alkanes, the paracors can be expressed by

\[ P_i = 25.2 + 2.86 M_i \]  

(3.137)  

Several authors propose paracors for pure hydrocarbons that devi-
ate from the group-contribution values. For example, \( P_{C_7} = 77 \) is
often cited for methane instead of the group-contribution value of
\( P_{C_7} = 71 \). Likewise, \( P_{C_8} = 41 \) is often used for nitrogen instead of
the group-contribution value of \( P_{C_8} = 35 \). Fig. 3.23 plots paracors
vs. molecular weight for pure components and petroleum fractions.
TABLE 3.2—PARACHORS FOR PURE COMPONENTS AND COMPOUND GROUPS

<table>
<thead>
<tr>
<th>Pure Component</th>
<th>parachor ( \gamma_i )</th>
</tr>
</thead>
<tbody>
<tr>
<td>( C_1 )</td>
<td>71</td>
</tr>
<tr>
<td>( C_2 )</td>
<td>111</td>
</tr>
<tr>
<td>( C_3 )</td>
<td>151</td>
</tr>
<tr>
<td>( C_4 )</td>
<td>191</td>
</tr>
<tr>
<td>( C_5 )</td>
<td>231</td>
</tr>
<tr>
<td>( C_6 )</td>
<td>271</td>
</tr>
<tr>
<td>( C_7 )</td>
<td>311</td>
</tr>
<tr>
<td>( C_8 )</td>
<td>351</td>
</tr>
<tr>
<td>( C_9 )</td>
<td>391</td>
</tr>
<tr>
<td>( C_{10} )</td>
<td>431</td>
</tr>
<tr>
<td>( N_2 )</td>
<td>35</td>
</tr>
<tr>
<td>( CO_2 )</td>
<td>49</td>
</tr>
<tr>
<td>( H_2S )</td>
<td>80</td>
</tr>
<tr>
<td>Group</td>
<td></td>
</tr>
<tr>
<td>( C )</td>
<td>9.0</td>
</tr>
<tr>
<td>( H )</td>
<td>15.5</td>
</tr>
<tr>
<td>( CH_3 )</td>
<td>55.5</td>
</tr>
<tr>
<td>( CH_2 )</td>
<td>40.0</td>
</tr>
<tr>
<td>( N )</td>
<td>17.5</td>
</tr>
<tr>
<td>( O )</td>
<td>20.0</td>
</tr>
<tr>
<td>( S )</td>
<td>49.1</td>
</tr>
</tbody>
</table>

Example: For methane, \( CH_4 \), \( \rho_{Me} \approx 0.74/(P_i/294) \) (4 + (15.5) - 71).

Nokay\(^{90}\) gave a simple relation for parachors of pure hydrocarbons (paraffins, olefins, naphthenes, and aromatics) with a normal boiling point between 400 and 1,400°F and specific gravity less than 1.

\[
\log P_i = -4.20895 + 2.29319 \log \left( \frac{T_i}{10} \right) \tag{3.138}
\]

with \( T_i \) in °R.

Katz and Saltman\(^{87}\) and Katz et al.\(^{88}\) give parachor data for \( C_7 \) fraction measured by Standing and Katz\(^{89,90}\) which are approximately correlated by

\[
P_i = 35 + 2.40M_i \tag{3.139}
\]

The API recommended procedure for estimating petroleum fraction IFT's is based on an unpublished correlation.\(^{27}\) The graphical correlation can be expressed by

\[
\sigma_i = 602(1 - T_i/314)^{0.194} \tag{3.140}
\]

where \( K \) is \( T_i^{1/2} \); with \( T_i \) in °R. The parachor can be estimated with the Macleod relation,

\[
P_i = \rho_i^{0.14} \left( \frac{M_i}{RT_i} \right)^{0.5} \tag{3.141}
\]

where \( \rho_i > \rho_o \) is assumed. The saturated-liquid density can be estimated, for example, with the Rackett\(^{33}\) equation.

\[
\rho_{\text{sat}} = \frac{M_i \rho_o}{RT_i} \left[ 1 - \left( \frac{T_i}{Z_\infty} \right)^{0.27} \right] \tag{3.142}
\]

where \( Z_\infty = Z_o = 0.291 - 0.08 \omega_i \) \( \tag{3.143} \)

and \( R \) = universal gas constant. The parachors predicted from Eqs. 3.140 through 3.143 are practically constant for a given petroleum fraction (i.e., the temperature effect cancels out).

GAS AND OIL PROPERTIES AND CORRELATIONS

Firoozabadi et al.\(^{90}\) give an equation that can be used to approximate the parachor of pure hydrocarbons from \( C_1 \) through \( C_{14} \) and for \( C_{14} \) fractions,

\[
P_i = 11.4 + 3.23M_i - 0.0022M_i^2 \tag{3.144}
\]

They also discuss the qualitative effect of asphaltenes on IFT and suggest that the parachor of asphaltic substances generally will not follow the relations of lighter \( C_7 \) fractions.

Ramey\(^{91}\) gives a method for estimating gas/oil IFT with black-oil PVT properties. We extend the method here to include the effect of solution oil/gas ratio, \( r_o \). Considering stock-tank oil and separator gas as the "components" (\( \tau \) and \( \theta \)) making up reservoir oil and gas, respectively, the Weinaug-Katz\(^{92}\) relation can be written

\[
\phi_{\text{mix}} = P_0 \left[ \frac{\rho_o}{M_o} - \gamma_o \frac{\rho_x}{M_x} \right] + P \left[ \frac{\rho_x}{M_x} - \gamma_x \frac{\rho_o}{M_o} \right] \tag{3.145a}
\]

where \( x = \frac{1}{1 + (7.52 \times 10^{-8})R_i/M_o^2 \gamma_o} \) \( \tag{3.145b} \)

\[
x_o = 1 - x, \quad \gamma_o = 1 + (7.52 \times 10^{-8})M_o^2 \gamma_o \rho_o \tag{3.145c}
\]

\[
y_x = 1 - y, \quad \gamma_x = 62.4 \gamma + 0.036y K_x \tag{3.145d}
\]

\[
\rho_o = 1, \quad \rho_x = 0.0932 \rho_x/\gamma x \tag{3.145f}
\]

\[
M_o = x_o M_x + x_p M_p \tag{3.145g}
\]

Fig. 3.23—Hydrocarbon parachors.
\[ M_f = \gamma_2 M_s + \gamma_2 M_T, \quad (3.145i) \]
\[ M_s = 6.0844 M_{API} - 5.9, \quad (3.145j) \]
\[ M_T = 28.97 M_{API}, \quad (3.145k) \]
\[ P_s = (2.376 + 0.01027 M_{API})/M_s, \quad (3.145l) \]
\[ P_T = 25.2 + 2.86 M_T, \quad (3.145m) \]

where \( P_s \) is the saturated pressure, \( M_{API} \) is the API gravity, and \( M_T \) is the molecular weight of the mixture.

**5.3.2 Diffusion Coefficients.** Molecular diffusion in multicomponent mixtures is a complex problem. The standard engineering approach uses an effective diffusion coefficient for Component \( i \) in a mixture, \( D_{ii} \), where \( D_{ii} \) can be calculated in one of two ways: (1) from binary diffusion coefficients and mixture composition or (2) from Component \( i \) properties and mixture viscosity. The first approach uses the Wilke\(^6\) formula to calculate \( D_{ii} \),

\[ D_{ii} = \frac{1 - \gamma_i}{\sum_{j \neq i} \frac{\gamma_j D_{ij}}{D_j}}, \quad (3.146) \]

where \( \gamma_i \) is the mole fraction of Component \( i \) and \( D_j \) is the binary diffusivity at the pressure and temperature of the mixture.

Sigmund\(^7\) correlates the effect of pressure and temperature on diffusion coefficients using a corresponding-states approach with reduced density,

\[ \frac{\rho_{ii} D_{ii}}{\rho_{ii}^T} = 0.99589 + 0.09601 \rho_{ii} - 0.22035 \rho_{ii}^2 + 0.003274 \rho_{ii}^3, \quad (3.147) \]

where \( \rho_{ii} \) is the reduced density at pressure and temperature \( T \), \( M_{ii} \) is the molecular weight of Component \( i \), and \( D_{ii}^T \) is the low-pressure density-diffusivity product.

dSa Silva and Belery\(^8\) note that the Sigmund correlation does not work well for liquid systems and propose the following extrapolation for \( \rho > 3 \),

\[ \frac{\rho_{ii} D_{ii}}{\rho_{ii}^T} = 0.18839 \exp(3 - \rho_{ii}), \quad (3.148) \]

which avoids negative \( D_{ii} \) for oils at \( \rho > 3 \) as estimated by the Sigmund correlation.

Low-pressure binary gas diffusion coefficients, \( D_{ii}^0 \), can be estimated from

\[ D_{ii}^0 = 0.001883 \frac{T^{0.5} \left[1/M_i + 1/M_j \right]^{0.5}}{p_{ii}^T \Omega_i}, \quad (3.149a) \]

where \( \Omega_i = \frac{1.0636}{T_i^{0.5}} + 0.193 \exp(0.47635 T_i) \]

\[ + 1.03587 \exp(1.52997 T_i) \quad + 1.76474 \exp(3.39411 T_i) \quad (3.149b) \]

\[ T_i = \frac{T}{(k/k_i)^{0.5}}, \quad (3.149c) \]

\[ (k/k_i)^{0.5} = \frac{[2.355 + 0.754 M_i^{0.5}]}{k}, \quad (3.149d) \]

\[ (k/k_i)^{0.5} = 65.3T_i^{0.5}, \quad (3.149e) \]

\[ \alpha_i = 0.5(\alpha_i + \alpha_j), \quad (3.149f) \]

\[ \alpha_j = 0.1866 \frac{v_{ii}^{0.5}}{v_{ii}^{0.5}}, \quad (3.149g) \]

with the diffusion coefficient, \( D_{ii}^0 \), in scf/STb, molecular weight, \( M_i \), in kg/mol, temperature, \( T \), in K, pressure, \( p \), in bar; Lennard Jones 12-6 potential parameter, \( \alpha \), in Å; and \( \alpha_i \) in m/mol and where \( Z_i \) is critical compressibility factor and \( \rho_{ii}^T \) is diffusing and concentrated species, respectively.

To obtain the low-pressure density-diffusivity product, we use the ideal gas law, \( p_{ii}^T = p_{ii}/RT_i \), to get

\[ D_{ii}^0 = \left(2.2648 \times 10^{-5}\right) \frac{T^{0.5} [1/M_i + 1/M_j]^{0.25}}{\alpha_i^{0.1} \Omega_i}, \quad (3.150) \]

where \( \rho \) and \( \rho_{ii}^T \) have units g mol/cm\(^3\).

The accuracy of the Sigmund correlation for liquids is not known, but the extension proposed by da Silva and Belery\(^8\) for large reduced densities does avoid negative diffusivities calculated by the Sigmund equation.\(^9\) Renner\(^10\) proposes a generalized correlation for effective diffusion coefficients of light hydrocarbons and CO\(_2\) in reservoir liquids that can be used as an alternative to the Sigmund-type correlation.

\[ D_{ii} = 10^{-1.8} \rho_{ii}^{-0.5} M_i^{1.08} p_{ii}^{0.2} T_i^{0.3} \exp(-1.83 I_{T,45}), \quad (3.151) \]

with \( D \) in cm\(^2\)/s and where \( \mu_i \) is viscosity in cp, \( M_i \) is molecular weight, \( \rho_{ii} \) is mole density of Component \( i \) at \( T \) and \( p \) in g mol/cm\(^3\), \( p \) is pressure in psia, and \( T \) is temperature in K. This correlation is based on 141 experimental data with the following property ranges: 0.2 < \( \mu_i < 134 \) cp; 16 < \( M_i < 44 \); 0.04 < \( \rho_{ii} < 10 \) kmol/m\(^3\); 14.7 < \( p < 2.560 \) psia; and 273 < \( T < 333 \) K. \( I_{T,45} \) is a function of CO\(_2\), C\(_1\), C\(_2\), and C\(_3\).

Renner also gives a correlation for diffusivity of CO\(_2\) in water/brine systems.

\[ D_{CO_2} = (6.392 \times 10^{-3}) \rho_{CO_2}^{0.27} M_i^{1.08} p_{ii}^{0.2} T_i^{0.3} \exp(-1.83 I_{T,45}), \quad (3.152) \]

with \( D \) in cm\(^2\)/s and \( \mu_i \) in cp.

**3.6 K-Value Correlations.** This section covers the estimation of equilibrium K values by correlations and the calculation of two-phase equilibrium when K values are known. The K value is defined as the ratio of equilibrium gas composition \( y_i \) to the equilibrium liquid composition \( x_i \).

\[ K_{i} = y_i/x_i, \quad (3.153) \]

\( K_i \) is a function of pressure, temperature, and overall composition. K values can be estimated with empirical correlations or by satisfying the equal-fugacity constraint with an EOS (see Chap. 4).

Although the increasing use of EOS’s has tended to lessen interest in empirical K-value correlations, empirical methods are still useful for such engineering calculations as (1) multistage surface separation, (2) compositional reservoir material balance, and (3) checking the consistency of separator-oil and gas compositions.

**PHASE BEHAVIOR**
Several methods for correlating $K$ values have appeared in the past 50 years. Most rely on two limiting conditions for describing the pressure dependence of $K$ values. First, at low pressures, Raoult’s and Dalton’s laws\(^3\) can be used to show that

$$K_i = \frac{p_v(T)}{p} \quad \text{...} \quad (3.154)$$

where $p_v$ = component vapor pressure at the system temperature. The limitations of this equation are that temperature must be less than the component critical temperature (because vapor pressure is not defined at supercritical temperatures) and that the component behaves as an ideal gas. Also, the equation implies that the $K$ value is independent of overall composition. In fact, the pressure dependence of low-pressure $K$ values is closely approximated by Eq. 3.154.

The second observation is that, at high pressures, the $K$ values of all components in a mixture tend to converge to unity at the same pressure. This pressure is called the convergence pressure\(^9\) and, for binaries, represents the actual mixture critical pressure. For multicomponent mixtures, the convergence pressure is a nonphysical condition unless the system temperature equals the mixture critical temperature.\(^9\) This is because a mixture becomes single phase at the bubblepoint or dewpoint pressure before reaching the convergence pressure.

The log-log plot of $K_i$ vs. pressure in Fig. 3.24 shows how the ideal-gas and convergence-pressure conditions define the $K$-value behavior at limiting conditions. For light components (where $T < T_{ci}$), $K$ values decrease monotonically toward the convergence pressure. For heavier components (where $T > T_{ci}$), $K$ values initially decrease as a function of pressure at low pressures, passing through unity when system pressure equals the vapor pressure of a particular component, reaching a minimum, and finally increasing toward unity at the convergence pressure.

For reservoir fluids, the pressure where $K$ values reach a minimum is usually $>1,000$ psia (Fig. 3.25), implying that $K$ values are more or less independent of convergence pressure (i.e., composition) at pressures $<1,000$ psia. This observation has been used to develop general “low-pressure” $K$-value correlations for surface-separator calculations.

### 3.6.1 Hoffman et al. Method

Hoffman et al.\(^9\) propose a method for correlating $K$ values that has received widespread application.

$$K_i = \frac{10^{[A_0 + A_1 F_i]}}{p} \quad \text{or} \quad \log K_i p = A_0 + A_1 F_i \quad \text{...} \quad (3.155)$$

where $F_i = \frac{1}{T_{ci}} - \frac{1}{T_{ci}} \log \left( \frac{p_v}{p_{ci}} \right) \quad \text{...} \quad (3.156)$

$T_c$ = critical temperature; $p_c$ = pressure; $T_b$ = normal boiling point; $p_{sc}$ = pressure at standard conditions; and $A_0$ and $A_1$ = slope and intercept, respectively, of the plot $\log(K_i p)$ vs. $F_i$.

Hoffman et al. show that measured $K$ values for a reservoir gas condensate correlate well with the proposed equation. They found that trend of $\log(K_i p)$ vs. $F_i$ is linear for components $C_1$ through $C_6$ at all pressures, while the function turns downward for heavier components at low pressures. Interestingly, the trend becomes more linear for all components at higher pressures. As Fig. 3.26 shows, Slope $A_1$ and Intercept $A_0$ vary with pressure. For low pressures, $K_i = p_v/p$. With the Clapeyron vapor pressure relation,\(^5\) $\log(p_v) = a - b/T$ results in $A_0 = \log(p_{vi})$ and $A_1 = 1$. These limiting values of $A_0$ and $A_1$ are close to the values found when $A_0(p)$ and $A_1(p)$ are extrapolated to $p = p_{ci}$. Because
$K$ values tend toward unity as pressure approaches the convergence pressure, $p_K$, it is necessary that $A_0 = \log(p_K)$ and $A_1 \to 0$. Several authors have noted that plots of $\log(K/p)$ vs. $F_i$ tend to converge at a common point. Brinkman and Sicking suggest that this “pivot” point represents the convergence pressure where $K_i = 1$ and $p = p_i$. The value of $F_i$ at the pivot point, $F_{Ki}$, is easily shown to equal $\log(p_K/p_i)$.

It is interesting to note that the well-known Wilson equation,

$$K_i = \exp \frac{5.37(1 + \omega_i)(1 - T_i^{-1})}{p_i},$$

(3.157)

is identical to the Hoffman et al. relation for $A_0 = \log(p_K)$ and $A_1 = 1$ when the Edmister correlation for acentric factor equation,

$$A_i = \frac{3}{7} \frac{T_i/T_{ca}}{1 - T_i/T_{ca}} \log(p_i/p_a) - 1,$$

(3.158)

is used in the Wilson equation. Note that $5.37 = (7/3) \ln(10)$.

Whitson and Torp suggest a generalized form of the Hoffman et al. equation in terms of convergence pressure and acentric factor.

$$K_i = \left[\frac{p_i}{p_K}\right]^{A_1-1} \exp \frac{5.37 A_1(1 + \omega_i)(1 - T_i^{-1})}{p_a},$$

(3.159)

where $A_1$ is a function of pressure, with $A_1 = 1$ at $p = p_a$ and $A_1 = 0$ at $p = p_K$. The key characteristics of $K$ values vs. pressure

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Fig. 3.26—Pressure dependence of slope, \( A_1 \), and intercept, \( A_0 \), in Hoffman et al. \( K_pF \) relationship (Eq. 3.155) for a North Sea gas condensate NS-1 (from Whitson and Torp[100]),

and temperature are correctly predicted by Eq. 3.159, where the following pressure dependence for \( A_1 \) is suggested.

\[
A_1 = 1 - \frac{1}{p/p_k^{3/2}} \quad \text{for } p < 1,000 \text{ psia and } T < 200 \text{°F}. \quad (3.160)
\]

where \( A_1 \) ranges from 0.5 to 0.8 and pressures \( p \) and \( p_k \) are given in psig. Canfield[105] also suggests a simple \( K \)-value correlation based on convergence pressure.

3.6.2 Standing Low-Pressure \( K \) Values. Standing[106] uses the Hoffman et al.[29] method to generate a low-pressure \( K \)-value equation for surface-separator calculations (when \( p < 1,000 \text{ psia} \) and \( T < 200 \text{°F} \)). Standing fits \( A_1 \) and \( A_0 \) in Eq. 3.155 as a function of pressure using \( K \)-value data from an Oklahoma City crude oil. He treats the \( C_7 \) fraction by correlating the behavior of \( K_{C_7} \) as a function of “effective” carbon number \( n_{C_{7+}} \). The Standing equations are

\[
K_i = \frac{F_i}{F_0} 10^{[A_0 + A_1/r_i]} \quad \text{for } i = C_7, \quad (3.161a)
\]

\[
F_i = \log_{p_i/p_k}/(1/T_i - 1/o_i) \quad \text{for } i = C_7, \quad (3.161b)
\]

\[
b_i = \log_{p_i/p_k}/(1/T_i - 1/o_i) \quad \text{for } i = C_7, \quad (3.161c)
\]

\[
A_{1i}(\rho) = 1.2 + \left[ (4.5 \times 10^{-5})p + (15 \times 10^{-5})p^2 \right], \quad (3.161d)
\]

\[
A_{1i}(\rho) = 0.890 - \left[ (1.7 \times 10^{-5})p + (3.5 \times 10^{-5})p^2 \right], \quad (3.161e)
\]

\[
n_{C_{7+}} = 7.3 + 0.0075\rho + 0.0016\rho^2 \quad \text{for } i = C_7, \quad (3.161f)
\]

\[
b_{i+} = 1.013 + 324n_{C_{7+}} - 4.256n_{C_{7+}}^2 \quad \text{for } i = C_7, \quad (3.161g)
\]

\[
T_{C_7} = 301 + 59.85n_{C_{7+}} - 0.971n_{C_{7+}}^2 \quad \text{for } i = C_7, \quad (3.161h)
\]

with \( \rho \) in °R except when calculating \( n_{C_{7+}} \) (for \( n_{C_{7+}}, T \) is in °F) and \( p \) in psia. Standing suggests modified values of \( b_i \) and \( T_{C_7} \) for nonhydrocarbons, methane, and ethane (Table 3.3). Glass et al. found that these equations are accurate for separator flash calculations of crude oils with GOR ranging from 300 to 1,500 scf/STB and oil gravity ranging from 26 to 48°API. Experience shows, however, that significant errors in calculated GOR may result for lean gas condensates, probably because of inaccurate \( C_1 \) and \( C_2 \) \( K \)-values. The Hoffman et al. method with Standing’s low-pressure correlations are particularly useful for checking the consistency of separator-gas and oil compositions.

3.6.3 Galimberti-Campbell Method. Galimberti and Campbell[110,111] suggested another useful approach for correlating \( K \)-values where

\[
\log K_i = A_0 + A_1 T_i^{2} \quad \text{for } i = C_7, \quad (3.162)
\]

is shown to correlate \( K \)-values for several simple mixtures containing hydrocarbons \( C_1 \) through \( C_{10} \) at pressures up to 3,000 psia and temperatures from −60 to 300°F. Whitson developed a low-pressure \( K \)-value correlation, based on data from Roland[110] at pressures < 1,000 psia and temperatures from 40 to 200°F, for separator calculations of gas condensates.

\[
A_0 = 4.276 - (7.6 \times 10^{-5})T + \left[ -1.18 + (5.675 \times 10^{-5})T \right] \log p, \quad \text{for } p < 1,000 \text{ psia} \quad (3.163a)
\]

\[
A_1 = 10^{-4} \left[ -4.9563 + 0.009557T \right] + \left[ 1.9904 \times 10^{-4} \right], \quad \text{for } p < 1,000 \text{ psia} \quad (3.163b)
\]

\[
T_{C_7} = 343 - 0.04p, \quad \text{for } p < 1,000 \text{ psia} \quad (3.163c)
\]

\[
tan \angle T_{C_7} = 1.0525 + 0.5125T + 0.00375T^2, \quad \text{for } p > 1,000 \text{ psia} \quad (3.163d)
\]

with \( p \) in psia, \( T \) in °F, and \( T \) in °R.

3.6.4 Nonhydrocarbon \( K \) Values. Lohrenz et al.[111] reported nonhydrocarbon \( K \)-values as a function of pressure, temperature, and convergence pressure.

\[
\ln K_{H_2S} = \left[ 1 - \frac{p}{F_T} \right]^{0.18} \quad 6.399217 + 1.389.2204 \frac{F_T}{T} \quad (3.164a)
\]

\[
-0.76885112 \ln p - 18.210525 \ln p \quad (3.164a)
\]

\[
-1.112 \frac{4.464}{T} \quad \text{for } p < 1,000 \text{ psia, } T < 200 \text{°F}, \text{ and } T < 200 \text{°F} \quad (3.164a)
\]

<table>
<thead>
<tr>
<th>Component, ( i )</th>
<th>( T_1 ) (cycle°R)</th>
<th>( T_2 ) (°R)</th>
</tr>
</thead>
<tbody>
<tr>
<td>( N_2 )</td>
<td>470</td>
<td>109</td>
</tr>
<tr>
<td>( CO_2 )</td>
<td>652</td>
<td>194</td>
</tr>
<tr>
<td>( H_2S )</td>
<td>1,136</td>
<td>331</td>
</tr>
<tr>
<td>( C_1 )</td>
<td>300</td>
<td>94</td>
</tr>
<tr>
<td>( C_2 )</td>
<td>1,145</td>
<td>303</td>
</tr>
<tr>
<td>( C_3 )</td>
<td>1,799</td>
<td>416</td>
</tr>
<tr>
<td>( i-C_4 )</td>
<td>2,037</td>
<td>471</td>
</tr>
<tr>
<td>( n-C_4 )</td>
<td>2,153</td>
<td>491</td>
</tr>
<tr>
<td>( i-C_5 )</td>
<td>2,368</td>
<td>542</td>
</tr>
<tr>
<td>( n-C_5 )</td>
<td>2,480</td>
<td>557</td>
</tr>
<tr>
<td>( C_6 ) (lumped)</td>
<td>2,738</td>
<td>610</td>
</tr>
<tr>
<td>( n-C_6 )</td>
<td>2,780</td>
<td>616</td>
</tr>
<tr>
<td>( n-C_7 )</td>
<td>3,068</td>
<td>669</td>
</tr>
<tr>
<td>( n-C_8 )</td>
<td>3,335</td>
<td>718</td>
</tr>
<tr>
<td>( n-C_9 )</td>
<td>3,590</td>
<td>763</td>
</tr>
<tr>
<td>( n-C_{10} )</td>
<td>3,828</td>
<td>805</td>
</tr>
</tbody>
</table>

For \( C_{1+} \) fractions, see Eqs. 3.161 through 3.161h.
In $K_{Ni} = 1 - \frac{p}{p_c}^{0.4} \frac{11.294748 - 1,184.2409}{f} - 0.90459907 \ln p$, \hspace{1cm} (3.164b)

In $K_{C_{Ni}} = 1 - \frac{p}{p_c}^{0.4} \left( \frac{7.020191 - 152.7291}{f} - 1.8896974 \times \ln p + 1.719,2956 \ln p - 644,740.69 \ln p \right)^{0.2}$, \hspace{1cm} (3.164c)

with $p$ in psia and $T$ in °R. For low-pressure $K$-value estimation, the first term in Eq. 3.164 simplifies to unity (assuming that $1 - p/p_c \approx 1$) and the $K$ values become functions of pressure and temperature only. However, these equations do not give the correct low-pressure value of $d(ln K_i)/d(ln p) = -1$

### 3.6.5 Convergence-Pressure Estimation.

Convergence pressure is used as a variable to define the composition dependence of $K$ values. Convergence pressure is a function of overall composition and temperature. Whiston and Michelsen\(^\text{112}\) show that convergence pressure is a thermodynamic phenomenon, with the characteristics of a true mixture critical point, that can be predicted with EOS's.

Bezas et al.\(^\text{113}\) give an empirical correlation for convergence pressure as a function of temperature and the product ($M_y \cdot e_y$).\(^{108}\) Standing\(^\text{2}\) suggests that convergence pressure of reservoir fluids varies almost linearly with $C_{Ni}$, molecular weight.

Convergence pressure can also be calculated with a trial-and-error procedure suggested by Rowe.\(^\text{97,98,114}\) This procedure involves the use of several empirical correlations for estimating mixture critical pressure and temperature, pseudocomponent critical properties, and the $K$ values of methane and octane. The Galimberti and Camp-berry\(^\text{108,109}\) $K$-value method is used to estimate $K$ values of other components by interpolation and extrapolation of the $C_1$ and $C_8$ $K$ values. This approach to convergence pressure is necessary if the $K$ values are used for processes that approach critical conditions or where $K$ values change significantly because of overall composition effects. The method cannot, of course, be more accurate than the correlations it uses and therefore is expected to yield only qualitatively correct results.

For reservoir calculations where convergence pressure can be assumed constant (e.g., pressure depletion), a more direct approach to determining convergence pressure is suggested. With a $K$-value correlation of the form $K_i = K(p_E, p_T)$ as in Eq. 3.159, the convergence pressure can be estimated from a single experimental saturation pressure. For a bubblepoint and a dewpoint, Eqs. 3.165 and 3.166, respectively, must be satisfied.

$$F[p_E] = 1 - \sum \frac{z_i}{K_i(p_E, p_T)} = 0 \hspace{1cm} (3.165)$$

$$and \hspace{1cm} P[p_T] = 1 - \sum \frac{z_i}{K_i(p_E, p_T)} = 0 \hspace{1cm} (3.166)$$

where $z_i$, $p_E$, or $p_T$ are specified and $p_D$ is determined.

The two-phase flash calculation with $K$ values given, is discussed in Chap. 4 in the Phase-Split Calculation section.

### References

1. The SI Metric System of Units and SPE Metric Standard, SPE, Richmond, Texas (June 1982).

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**SI Metric Conversion Factors**

\[
\begin{align*}
\text{Å} & \times 1.0^* & \text{E} & = 1.0 \times 1.0^* & \text{mm} & \\
^\circ\text{API} & 141.5/(131.5 + ^\circ\text{API}) & \text{g/cm}^3 \\
\text{bar} & \times 1.0^* & \text{E} & = 1.0 \times 1.0^* & \text{Pa} \\
\text{bbl} & \times 1.589 873 & \text{E} & = 1.0 \times 1.0^* & \text{m}^3 \\
\text{Btu/lbm mol} & \times 2.236 & \text{E} & = 1.0 \times 1.0^* & \text{J/mol} \\
\text{cp} & \times 1.0^* & \text{E} & = 1.0 \times 1.0^* & \text{s} \\
\text{cSt} & \times 1.0^* & \text{E} & = 1.0 \times 1.0^* & \text{m}^2/\text{s} \\
\text{ft} & \times 3.048 & \text{E} & = 1.0 \times 1.0^* & \text{m} \\
\text{ft}^2 & \times 0.290 304 & \text{E} & = 1.0 \times 1.0^* & \text{m}^2 \\
\text{ft}^3 & \times 1.341 817 & \text{E} & = 1.0 \times 1.0^* & \text{m}^3 \\
\text{ft}^3/\text{lbm mol} & \times 6.242 796 & \text{E} & = 1.0 \times 1.0^* & \text{m}^3/\text{kmol} \\
{^\circ}\text{F} & (^\circ\text{F} - 32)/1.8 & \text{K} \\
{^\circ}\text{F} & (^\circ\text{F} + 459.67)/1.8 & \text{K} \\
ie & \times 6.451 6^* & \text{E} & = 1.0 \times 1.0^* & \text{cm}^2 \\
\text{lbm} & \times 4.535 924 & \text{E} & = 1.0 \times 1.0^* & \text{kg} \\
\text{lbm mol} & \times 4.535 924 & \text{E} & = 1.0 \times 1.0^* & \text{kmol} \\
\text{psi} & \times 6.894 757 & \text{E} & = 1.0 \times 1.0^* & \text{MPa} \\
\text{psi}^{-1} & \times 1.450 377 & \text{E} & = 1.0 \times 1.0^* & \text{KPa}^{-1} \\
\text{R} & \times 5^9 & \text{E} & = 1.0 \times 1.0^* & \text{K} \\
\end{align*}
\]

*Conversion factor is exact.*