

## 9. RESERVOIR FLUID CALCULATIONS – LIQUID SYSTEMS

The PVT relation of liquids also is important in calculations pertaining to petroleum reservoir behavior. The isothermal compressibility of undersaturated liquid, the shrinkage in volume that a barrel of reservoir liquid undergoes while progressing from the reservoir to the Stock Tank, and the in-place density of the reservoir liquid are three examples that require a knowledge of hydrocarbon liquid behavior.

### 9.1. Liquid density

A liquid is a fluid without independent shape but having a definite volume for a fixed mass under given conditions. An ideal liquid is defined mathematically by the following partial derivatives:

$$\left(\frac{dV}{dT}\right)_P = 0 \text{ and } \left(\frac{dV}{dP}\right)_T = 0 \quad (9.1)$$

where:

$\left(\frac{dV}{dT}\right)_P$  is the isobaric thermal expansion

$\left(\frac{dV}{dP}\right)_T$  is the isothermal compressibility.

Coefficient of isothermal compressibility is expressed by :

$$c = -\frac{1}{V} \left(\frac{dV}{dP}\right)_T \quad (9.2)$$

Coefficient of isobaric thermal expansion:

$$\beta = \frac{1}{V} \left(\frac{dV}{dT}\right)_P \quad (9.3)$$

The relations state that changes in volume with pressure and temperature are zero for an ideal liquid. Furthermore, an ideal liquid is conceived to have no internal friction between molecules (i. e.  $\mu L = 0$ ). Thus, the concept of an ideal liquid is of little quantitative value in predicting the volumetric behavior.

There are several methods of calculating the volume that a given mass of liquid occupies under an elevated pressure and temperature. When the composition of a reservoir liquid is known, its volume at any pressure and temperature can be calculated either by applying ideal solution principles or by using partial volume methods. When the composition of the liquid is unknown, but the properties and quantities of surface gas and oil derived from the reservoir liquid are known, empirical methods are available to calculate the liquid volume under reservoir conditions.

We will outline first the methods that can be used when composition of the liquid is known. Following this, empirical methods will be presented that are based on surface gas-oil ratio measurements.

#### 9.1.1 Liquid Density from Composition Using Ideal Solution Principles

The behavior of mixtures of liquid hydrocarbons at atmospheric pressure and temperature approaches the behavior of ideal solutions. That is, a unit volume of one liquid, such as pentane, added to a unit volume of another liquid, such as heptane, will result in two volumes of mixture. This ideal solution behavior makes it possible to calculate the volume of a liquid mixture at 14.7 psia and 60°F from the composition of the liquid and the densities of the individual components. The atmospheric volume of the mixture can then be corrected for compressibility and thermal expansion to obtain its volume at an elevated pressure and temperature.

The density of stock-tank oil can be determined experimentally by means of a PAAR densitometer or by calculations based upon chemical component analysis using the additive volume technique.

Based on 1 mole liquid:

$$\rho_{STO} = \frac{m}{v} = \frac{\sum x_i M_i}{\sum \frac{x_i M_i}{\rho_i}} = \frac{M_{STO}}{\sum \frac{x_i M_i}{\rho_i}} \quad (9.4)$$

where:

$\rho_{STO}$ : density of STO at SC.

$m$ : mass of 1 mole (molar mass)

$v$ : volume of 1 mole (molar volume)

$x_i$ : mole fraction of individual components

$M_i$ : molecular weight of individual components

$\rho_i$ : density of individual components from handbook

$M_{STO}$ : molecular weight of STO

Usually, the content of light components as methane, ethane and propane is low, and the apparent liquid density values found in handbooks are used for these compounds.

### 9.1.2. Liquid Density from Composition analyses

Accurate determinations of liquid densities of live oil at different P and T are a serious problem in the petroleum literature. A technique very often applied in petroleum is to suppose an apparent liquid at standard conditions. An apparent density value of this liquid at standard conditions is then calculated. This value is then corrected for actual T and P.

It is, however, experimentally observed that the apparent liquid density of the two light hydrocarbons, methane and ethane, changes with the density of the total system. The apparent liquid density of the two compounds increases as the total density of the oil increases. Heavier molecules have a greater force of attraction. Thus, it is difficult to use fixed apparent density values of methane and ethane at standard conditions.

When applying the ideal solution method to reservoir liquid systems that contain large quantities of dissolved gases, it is apparent immediately that it is impossible for such systems to continue to be all liquid when brought to atmospheric pressure and temperature. However, this physical limitation does not impair the mathematical use of atmospheric liquid densities of methane and ethane in an ideal solution calculation, because the "pseudoliquid density" of the system at atmospheric conditions is only a stepping stone to the evaluation of the density at an elevated pressure and temperature.

Standing and Katz [1] studied measured data on systems containing methane plus other compounds and ethane plus other compounds. By taking the experimentally determined system density at numerous elevated pressures and temperatures and correcting this density to 14.7 psia and 60°F, a so-called pseudo-liquid density of the system was calculated. Subtracting the volume and weight contributed by the "plus" compound from the pseudo-density left an "apparent liquid density" of the methane or ethane. In other words, the apparent liquid densities of methane and ethane are fictitious densities the values which, used in an ideal solution calculation with the true liquid densities of other compounds, give an atmospheric system density that can be corrected to elevated pressures and temperatures by suitable compressibility and thermal expansion factors.

For convenience, the apparent liquid densities are calculated at conditions of 14.7 psia and 60°F. When the above procedure was carried out, the apparent liquid densities were found to vary with the density of the system as a whole, as shown in Fig. 9.1.

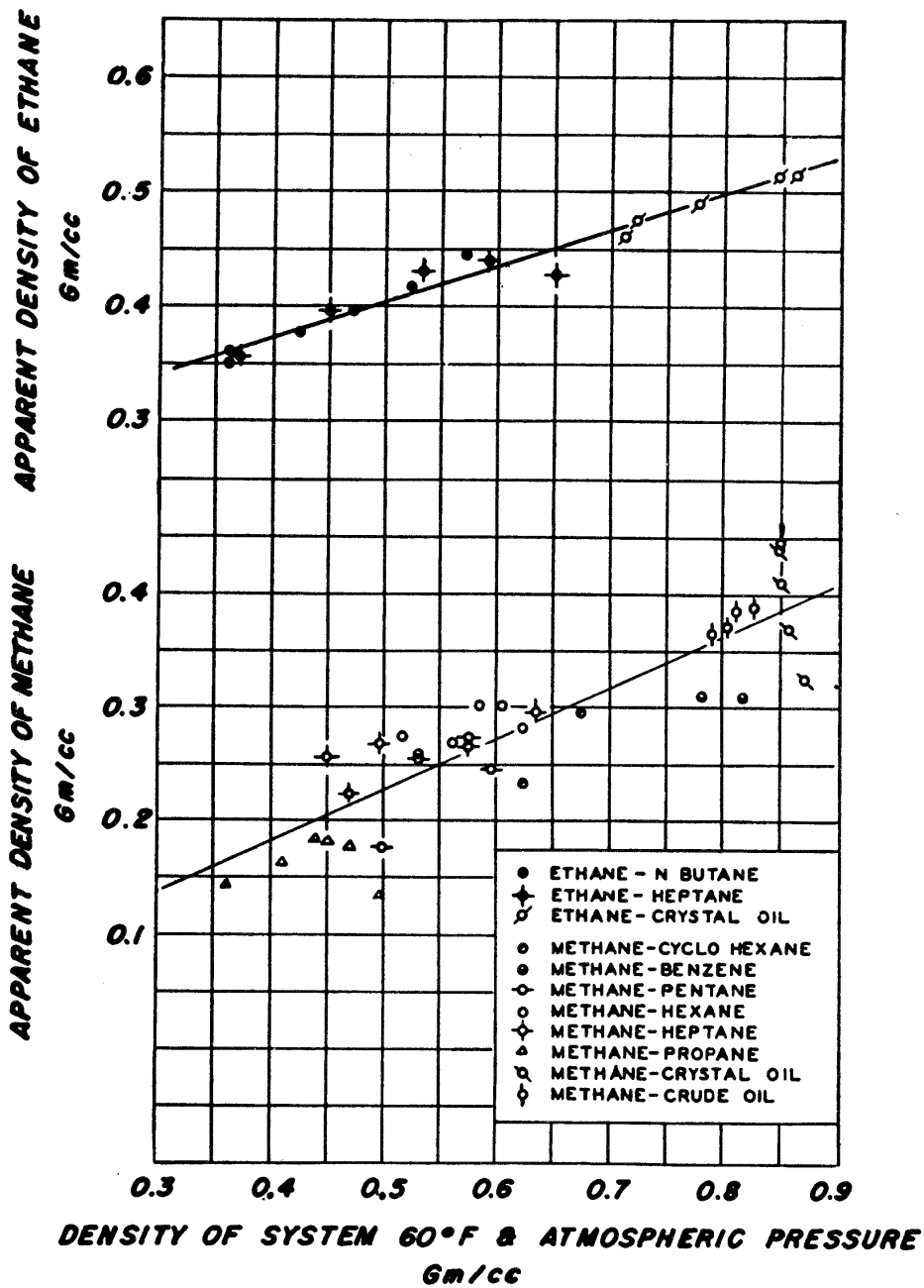


Fig.9.1. Relation of Apparent Density of Methane and Ethane to Density of the System  
(After Standing and Katz, Trans. AIME, 1942)

Direct use of the data in this form would require a trial-and-error solution; therefore, pseudo-liquid densities of the system have been converted into terms of the density of the heavier portion of the system and the weight % of the light component. The results of these conversions are presented in the form of the calculating chart shown in Fig. 9.2.

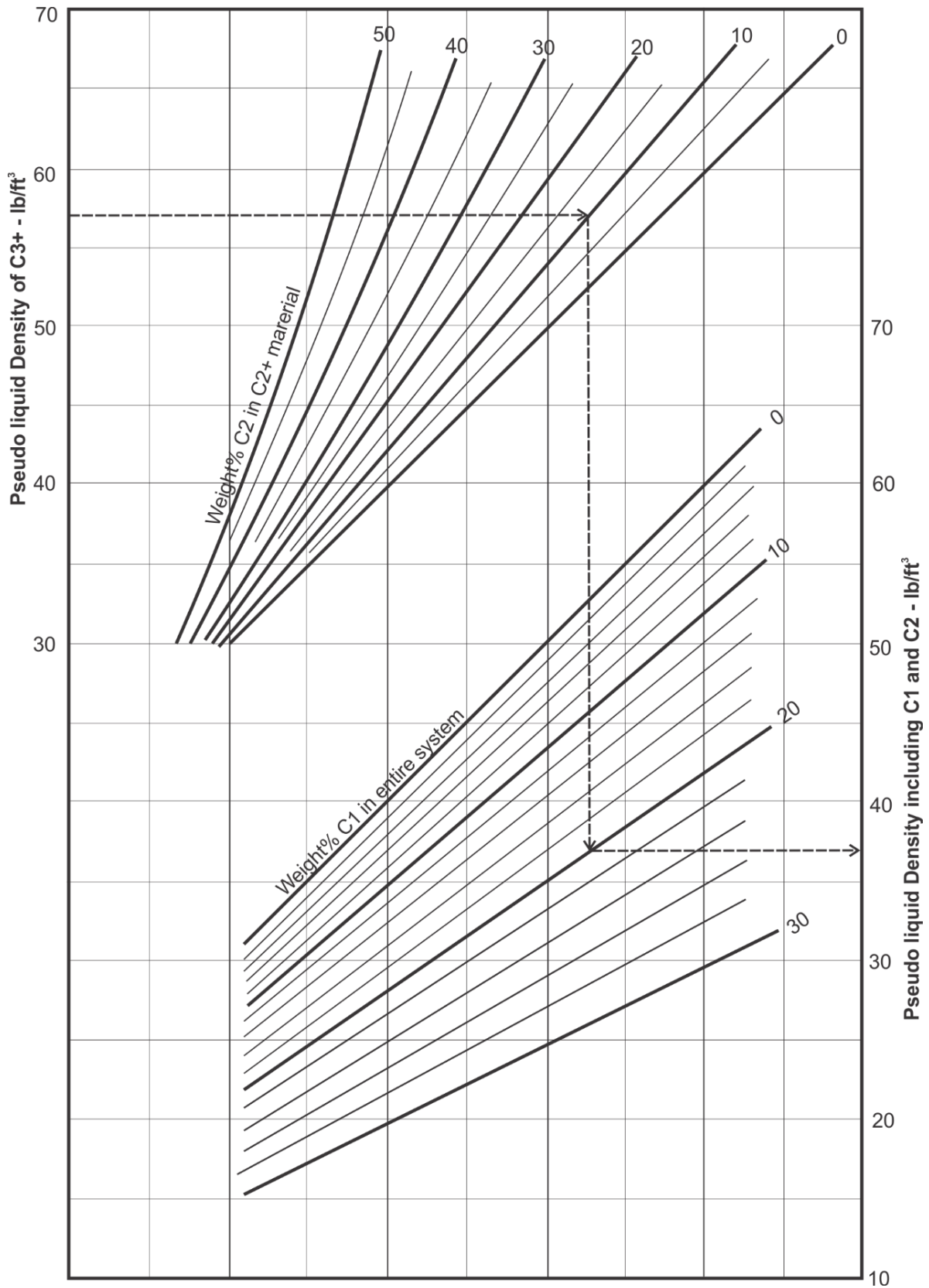


Fig. 9.2. Pseudo Liquid Density of Systems Containing Methane and Ethane at SC

To use Fig. 9.2, it is necessary to calculate the density of the propane plus fraction at 14.7 psia and 60°F, the weight% ethane in the ethane plus material, and the weight% methane in the entire system.

The formulas needed are based on 1 mole liquid, and mole fractions:  $x_1 = C1$ ,  $x_2 = C2$ ,  $x_3 = C3$ , etc. and components:

The density of the propane plus fraction at 14.7 psia and 60°F:

	$\rho_{C3+} = \frac{\sum_{i=3}^n (x_i M_i)}{\sum_{i=3}^n \left( \frac{x_i M_i}{\rho_i} \right)}$	(9.5)
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Wt% C2 in C2+fraction:

	$Wt\% C2 = \frac{100x_2 M_2}{\sum_{i=2}^n (x_i M_i)}$	(9.6)
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Wt% C1 in C2+fraction:

	$Wt\% C1 = \frac{100x_1 M_1}{\sum_{i=1}^n (x_i M_i)}$	(9.7)
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From these three values the pseudo-liquid density of the system at 14.7 psia and 60°F is determined directly from the chart. This procedure is illustrated in Example 9.1.

Temperature-specific gravity corrections for liquid hydrocarbons having specific gravities in the range 0.60-1.0 are given in the National Standard Petroleum Oil Tables [2]. The NGAA Standard Correction Factors [3] cover a lower temperature range than the NSPO Tables, in addition to lower gravities. Sage, Lacey, and co-workers on API Project 37 have measured the effect of temperature and pressure on the density for a number of multicomponent liquids. A survey of these data indicates:

- (1) The compressibility of hydrocarbon liquids at temperatures below 300°F can be assumed to be a function of the liquid density at 60°F. Minor differences exist in the liquid compressibility between systems of the same density, but the differences are not regular enough to take into account.
- (2) The thermal expansion of hydrocarbon liquids is affected little by pressure. Thus, it is possible to use the temperature corrections listed in the National Standard Petroleum Oil Tables" at elevated pressures, as well as at atmospheric pressure.

An additive correction factor to the atmospheric density of a liquid for the effect of pressure has been presented by Standing and Katz. [1] The range of the factor was later extended by Hanson, Kuist, and Brown [4]. This correction factor, presented in Fig. 9.3, makes it possible to correct the pseudoliquid densities to an elevated pressure state. For example, the liquid, whose pseudo-liquid density was calculated to be 50.8 lb/ft<sup>3</sup> at 14.7 psia and 60°F in Example 9.1, would have a density of 50.8 + 0.89 = 51.7 lb/ft<sup>3</sup> at 3280 psia and 60°F.

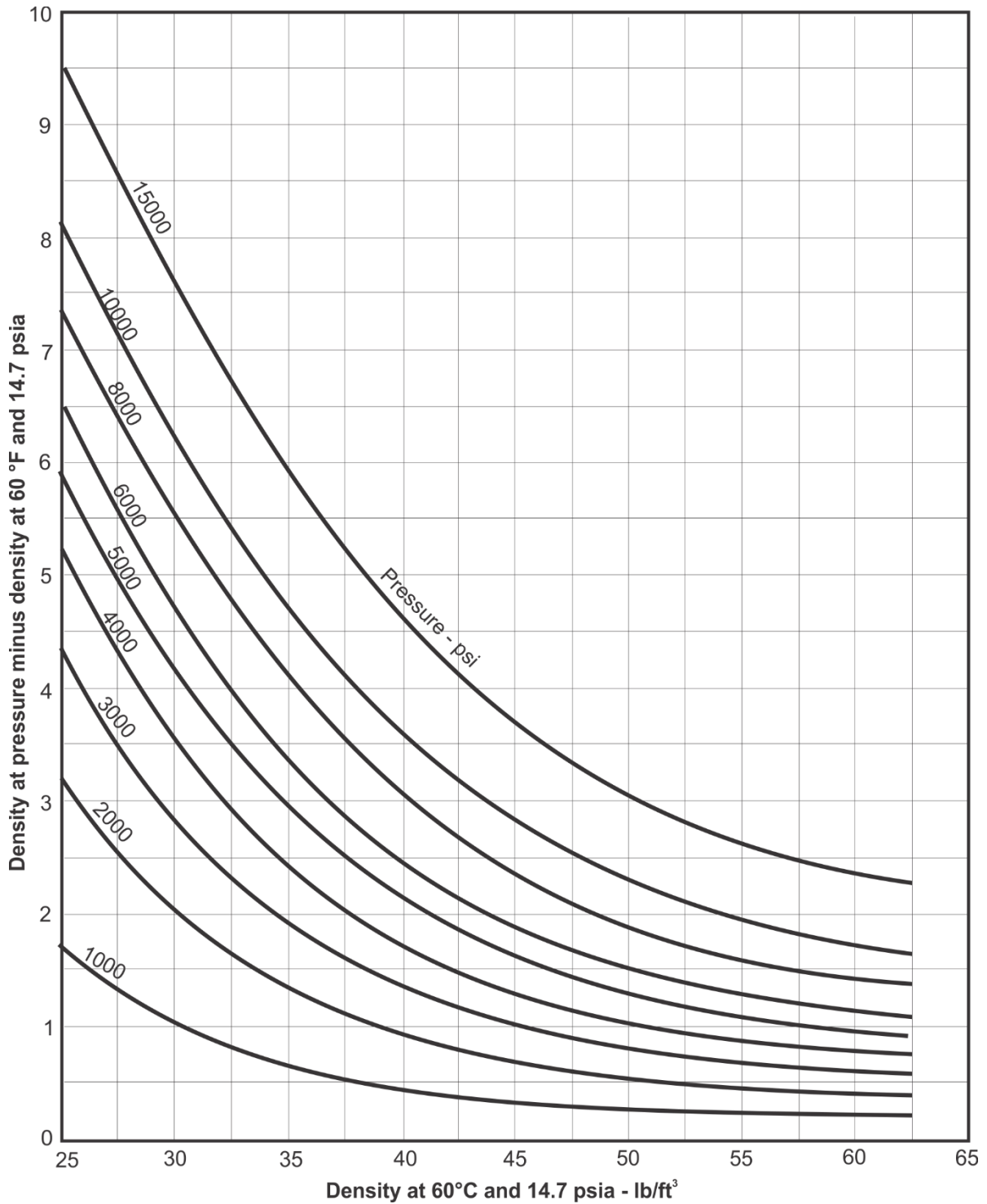


Fig. 9.3. Density Correction for Compressibility of Liquids.

Curve fit program for the pressure correction have been prepared by Standing:

$$\Delta\rho_P = \left\{ 0.0133 + 152.4(\rho_{app} + \Delta\rho_T)^{-2.45} \right\} (T - 60)^2 - \left\{ 8.1 \cdot 10^{-6} - 0.0622 \cdot 10^{-0.0764(\rho_{app} + \Delta\rho_T)} \right\} (T - 60)^2$$

where

$\rho_{app}$  = pseudo-liquid density at standard conditions (SC).

$\rho$  in lb/ft<sup>3</sup>

T in °F

The correction for the effect of temperature change is presented in Fig. 9.4.

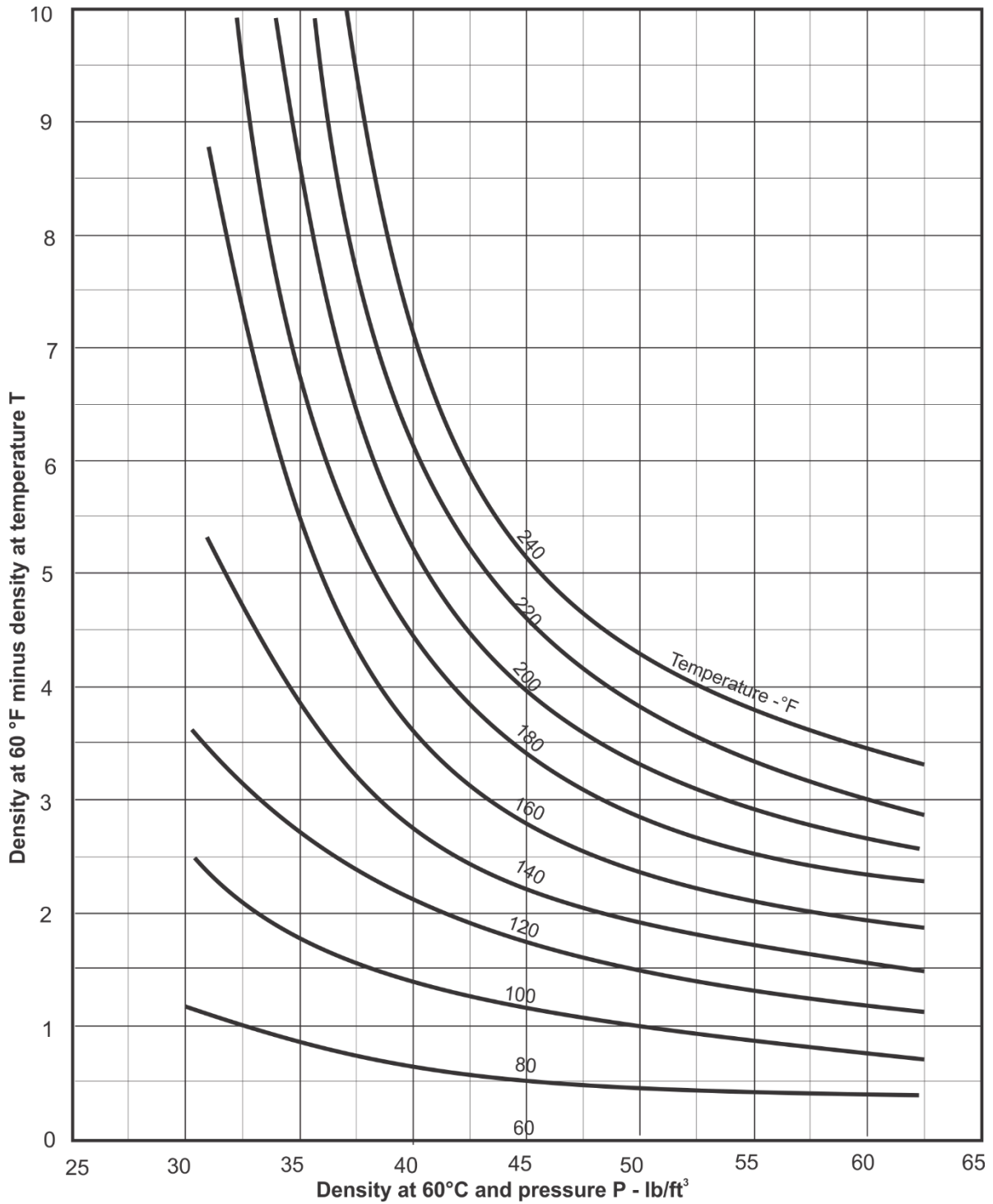


Fig. 9.4. Density Correction for Thermal Expansion of Liquids

Curve fit program for the temperature correction have been prepared by Standing:

$$\Delta\rho_T = (0.167 + 16.181 \cdot 10^{-0.0425\rho_{app}}) \left(\frac{P}{1000}\right) - 0.01(0.299 + 263 \cdot 10^{-0.0603\rho_{app}}) \left(\frac{P}{1000}\right)^2$$

where :

$\rho_{app}$  = pseudo-liquid density at standard conditions (SC).

$\rho$  = lb/ft<sup>3</sup>

$P$  = psia

These factors were obtained from the data of National Petroleum Tables, the NGA Standard Volume Correction Factors, and the data on low molecular weight liquids presented by Hanson, Kuist, and Brown [4].

Applying the correction factors of Fig. 9.4 to the Example 9.1 indicates that a correction factor of  $-3.57 \text{ lb/ft}^3$  should be applied to the 3280 psia-60°F density value to account for the thermal expansion of the compressed liquid in going from 60°F to 218°F. The density at 3280 psia and 218°F is, therefore,  $51.7 - 3.57 = 48.1 \text{ lb/ft}^3$ .

The method just outlined and illustrated in Example 9.1. usually will give densities less than 3 % in error. Standing and Katz [1] compared experimentally determined densities of fifteen liquids saturated with gas at pressures up to 8200 psi with calculated density values and found an arithmetic average error of only 1.16 %. The greatest error was 4 %. Checks with various data of Sage and Lacey have indicated comparable accuracy.

<b>Example 9.1.</b>					
To calculate the density of a crude oil-natural gas system at its bubble point of 3280 psia at 218°F, given the composition of the system using Ideal Solution Principles and Apparent Methane and Ethane Densities					
<b>Solution</b>					
Basis; One-pound mole.					
1	2	3	6	5	6
Comp.	Mole Frac. $z$	Molecular Weight $M$	Weight $m$ $z \cdot M$ , lb	Liquid Density $\rho_L$ 60°F/14.7 psia lb/ft <sup>3</sup>	Liquid Volume V <sub>L</sub> 60°F 14.7 psia $\frac{z \cdot M}{\rho}$ , ft <sup>3</sup>
C1	0.4404	16.0	7.046		
C2	0.0432	30.1	1.296		
C3	0.0405	44.1	1.786	31.66	0.0564
C4	0.0284	58.1	1.650	35.87*	0.0460
C5	0.0174	72.2	1.256	39.16*	0.0321
C6	0.0290	86.2	2.500	41.43	0.0603
C7+	0.4011	287	115.1	56.6	2.032
	1.0000		130.69		2.227

\*Arithmetic average of iso and normal values, Tbl. 8.1.- Gas Chapter.

Density propane plus

$$\rho_{C3+} = \frac{122.34}{2.227} = 54.93 \text{ lb/ft}^3$$

Weight% ethane in ethane plus

$$Wt\%C2 = \frac{1.296}{123.64} = 1.05 = 1.05 \text{ wt}\%$$

Weight% methane in system

$$Wt\%C1 = \frac{7.05}{130.69} = 5.4 \text{ Wt}\%$$

From Fig. 9.2: Pseudo-liquid density =  $50.8 \text{ lb/ft}^3$  at 60°F and 14.7 psia

From Fig. 9.3: Pressure correction to 3280 psia =  $+0.89 \text{ lb/ft}^3$

Density at 3280 psia and 60°F =  $50.8 + 0.89 = 51.7 \text{ lb/ft}^3$

From Fig. 9.4: Temperature correction to 218°F =  $-3.57 \text{ lb/ft}^3$

Density at 3280 psia and 218°F =  $51.7 - 3.57 = 48.1 \text{ lb/ft}^3$



### 9.1.3. Liquid Density from Composition and Partial Liquid Volumes

The specific volume of a liquid (reciprocal of density) at an elevated pressure and temperature state can be calculated by summing the products of the individual component weight fractions and their partial liquid volumes. This can be expressed mathematically as

$$\bar{V} = \bar{V}_{\sim 1} \cdot n_{\sim 1} + \bar{V}_{\sim 2} \cdot n_{\sim 2} + \dots + \bar{V}_{\sim j} \cdot n_{\sim j} \quad (9.8)$$

As in the gas calculations, the partial volumes of the individual components are a function of pressure, temperature, and the kind of material comprising the rest of the system. This third variable, in the case of hydrocarbon liquids found in petroleum reservoirs, is expressed by the average molecular weight of the other components in the system. Therefore, in using the partial volume method, it is necessary to evaluate both the weight fraction of every component in the system and the average molecular weight of all other components except the one under consideration. It can be appreciated that with three variables to consider, extensive cross-plotting of the data often is required.

Calculation of the specific volume of a liquid at 3280 psia and 218°F is shown in Example 9.2.

#### Example 9.2.

Calculation of the density of a crude oil-natural gas system at its bubble point 3280 psia at 218°F by the Method of Partial Volumes, given the composition of the system.

#### Solution:

Basis: One pound mole.

1	2	3	4	5	6	7	8
Comp.	Mole frac. n	Molecular Weight M	Weight n · M, lb	Weight frac. $\bar{n}$	Average M of Other Comp.	Partial Volume ft <sup>3</sup> /lb $\bar{V}_{\sim}$	Liquid Volume of Comp. 3280psia, 218°F $\bar{V}_{\sim} \cdot \bar{n}$
C1	0.4404	16.0	7.046	0.0539	221	0.09	0.00485
C2	0.0432	30.1	1.296	0.0099	135	0.0347	0.00034
C3	0.0405	44.1	1.786	0.0137	134	0.0347	0.00048
C4	0.0284	58.1	1.650	0.0126	133	0.0303	0.00038
C5	0.0174	72.2	1.256	0.0096	131	0.0252	0.00024
C6	0.0290	86.2	2.500	0.0191	23	0.0175	0.01575
C7+	0.4011	287	115.1	0.8812			
	1.0000		130.69	1.0000			0.02204

Calculated density of liquid at 3280 psia and 218° F

$$\rho = \frac{1}{0.02204} = 45.3 \text{ lb/ft}^3$$

#### Notes on Calculation Method

- (1) Density of C7+ fraction = 56.6 lb/ft<sup>3</sup> at 14.7 PSIA and 60° F
- (2) Partial volumes of methane through pentanes obtained from Sage, Hicks, and Lacey [5] Methane value can be greatly in error because of limited data.
- (3) Partial volume of C6+ at 3280 and 218° F obtained by use of Fig. 1,3 and Fig. 9.4 and the atmospheric density of C6+ (56.1 lb/ft<sup>3</sup>) to give a specific volume of 0.0186 ft<sup>3</sup>/lb. Partial volume of 0.0175 then obtained from use of 0.0011 ft<sup>3</sup>/lb residual volume and equation :

$$\bar{V}_{\sim} = \bar{V} - \bar{V}_{\sim}$$

The partial liquid volume data of methane through pentane have been interpolated from the tabular data of Sage, Hicks, and Lacey[5]. The specific volume of the hexanes plus fraction at 14.7 psia and 60°F has been corrected by the data of Fig. 9.3 and Fig. 9.4 to give the specific volume of this fraction at the elevated pressure and temperature. The partial volume of the hexanes plus was then obtained by subtracting a residual volume (tabulated by Sage, Hicks, and Lacey [5]) from the specific volume.

The accuracy of liquid density calculation made by the partial volume method is no better than that obtained by the empirical methods. Furthermore, the partial volume calculations are much more difficult to make. For these reasons, the partial volume method seldom is used. However, as the fund of partial volume data grows, the accuracy with which the density of liquids can be calculated will be increased. This probably is not true of the empirical methods.

9.1.4. Liquid Density from Production data

The first section of this chapter has shown how methane and ethane assume different apparent liquid densities when in different composition systems. Katz, [6] in developing a method of predicting the shrinkage of crude oils, extended the apparent density concept from methane and ethane to cover gases in general. Katz's method has the advantage that it does not require the composition of the system, but can be used where only the gas-oil ratio, gas gravity, and tank oil gravity are known. The method is almost as accurate as the method using the composition of the system and normally will give the liquid densities to within 3%.

The correlation found by Katz of the apparent density of dissolved gases as a function of gas gravity and tank oil gravity is given in Fig. 9.5.

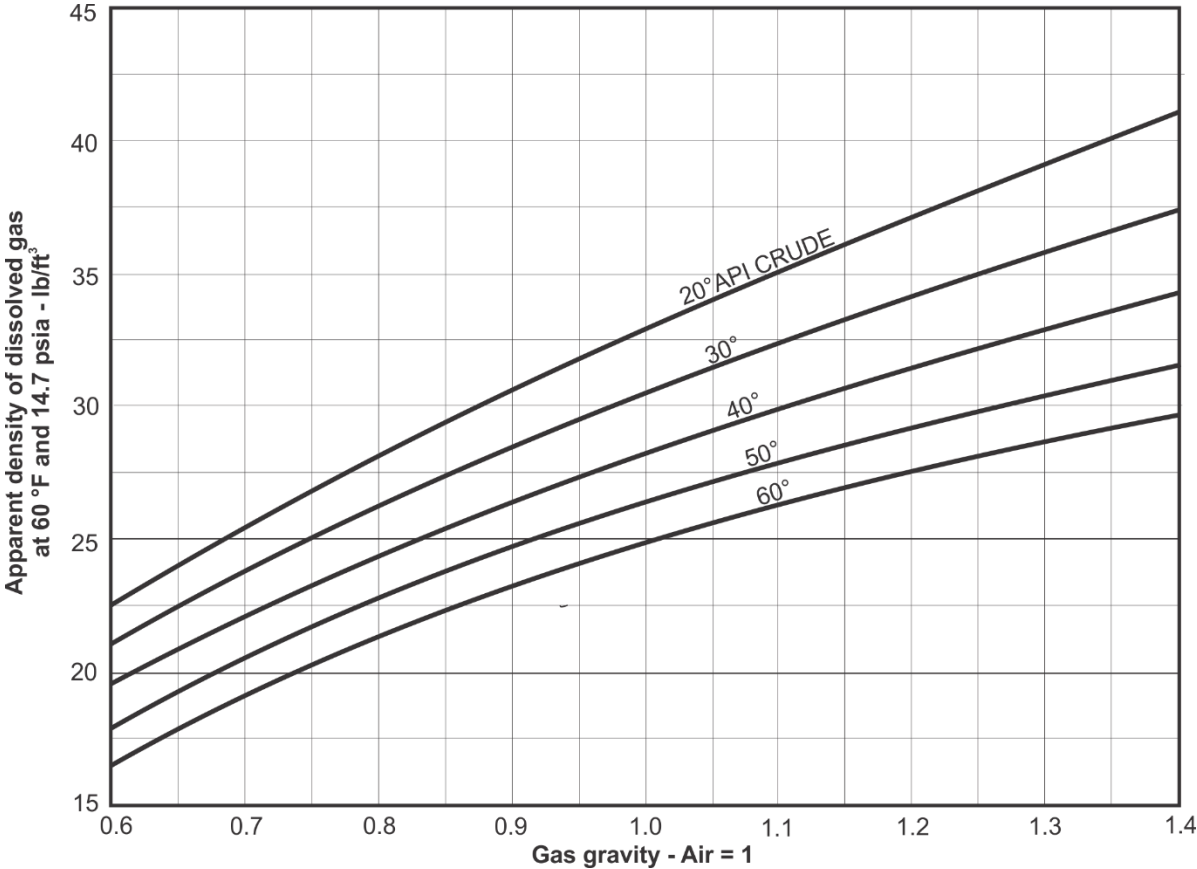


Fig. 9.5. Apparent Liquid Densities of Natural Gases [6] (After Katz, API Drilling and Production Practice, 1942)

An example of its use is worked out in Example 9.3, using gas gravities, tank oil gravity, and gas-oil ratio data to calculate the shrinkage, and a formation volume factor of the liquid used in the previous illustrations. The calculated formation factor of 1.279 bbl/bbl STO of tank oil compares favorably with the experimentally determined value of 1.302 bbl/bbl STO.

**Example9.3.**

Calculation of Shrinkage and Formation Volume Factor of a Crude Oil-Natural Gas Mixture Using Apparent Gas Density Method

**Problem:** To calculate the shrinkage that a reservoir liquid undergoes in being produced to the lease tank, given the surface production of gas and oil, the reservoir pressure and temperature, and assuming the liquid in the reservoir to be at its bubble point.

Given:

	Primary trap	Secondary trap	Tank (STO)	Total
Produced gas-oil ratio, ft <sup>3</sup> /bbl STO	414	90	25	529
Produced gas gravity, air = 1	0.640	0.897	1.54	

Tank oil gravity 27.4 °API = 0.891 sp.gr

**Solution:**

Average produced gas gravity,  $\gamma_{g,avg}$  :

$$\gamma_{g,avg} = \frac{414 \cdot 0.640 + 90 \cdot 0.897 + 25 \cdot 1.54}{414 + 90 + 25} = \frac{384.2}{529} = 0.726$$

Weight dissolved gas,  $m_g$ :

$$m_g = \frac{529}{379} \cdot (0.726 \cdot 28.97) = 29.3 \text{ lb}$$

Weight tank oil,  $m_{STO}$ :

$$m_{STO} = 0.891 \cdot 5.615 \cdot 62.4 = 312 \text{ lb}$$

From fig. 9.5;

Apparent density of dissolved gas = 24.9 lb/ft<sup>3</sup> at 14.7 psia and 60°F

$$\text{Pseudo - liquid volume of dissolved gas} = \frac{29.3}{24.9} = 1.177 \text{ ft}^3$$

	Volume, ft <sup>3</sup>	Weight, lb
Dissolved gas	1.177	29.3
Tank oil	5.615	312
Total	6.792	341.3

$$\text{Pseudo - liquid density} = \frac{341.2}{6.792} = 50.25 \text{ lb/ft}^3 \text{ at 14.7 psia and 60°F}$$

From Fig. 9.3; pressure correction to 3280 psia:

$$(\Delta\rho)_{\text{corrP}} = + 0.90 \text{ lb/ft}^3$$

Density at 3280 psia and 60°F

$$(\rho)_{\text{corrP}} = 50.25 + 0.90 = 51.15 \text{ lb/ft}^3$$

From Fig.9.4; temperature correction to 218°F

$$(\Delta\rho)_{\text{corrT}} = - 3.63 \text{ lb/ft}^3$$

Oil density at 3280 psia and 218°F

$$(\rho_o)_{\text{corrP\&T}} = 51.15 - 3.63 = 47.52 \text{ lb/ft}^3$$

Volume occupied at 3280 psia and 218°F (bubble point)

$$(V)_{bp} = \frac{341.3}{47.52} = 7.180 \text{ ft}^3$$

$$\text{Volume Shrinkage} = \frac{7.180 - 5.615}{5.615} 100 = 27.5\%$$

Formation volume factor at bp,  $(B_o)_{bp}$ :

$$(B_o)_{bp} = \frac{7.180}{5.615} = 1.279 \text{ bbl/bbl STO}$$

### 9.1.5 Liquid Density from Generalized Data of California Crudes

A second empirical method that can be used to calculate a liquid density employs a correlation developed by Standing [7] for California crude oils and natural gases. This empirical correlation of the formation volume of a natural gas-crude oil system at its bubble point condition is shown in Fig. 9.6 as a function of gas-oil ratio of the bubble point liquid,  $r_B$ ; the dissolved gas gravity,  $\gamma_g$ ; the tank oil specific gravity,  $\gamma_L$ ; and temperature,  $t$ .

The pressure is not used directly in this correlation, as it is an implicit function of the other variables. The formation volumes correlated in Fig. 9.6 checked, with an error less than 0.5%, 45% of the 105 bubble point liquids comprising the correlation. The arithmetic average of all errors was 1.17%. Although extensive checks have not been made with crudes from outside California, such crudes should not yield errors appreciably larger than those indicated above.

It will be noticed in the example shown in Example 9.3 that the calculation of the density and weight of a crude oil-natural gas system at its bubble point allows one to evaluate the formation volume at that condition. This relation can be used in reverse direction to calculate the density of a liquid from the weight of material in the system and the formation volume of the liquid at the desired pressure and temperature. An example of this method is shown in Example 9.4.

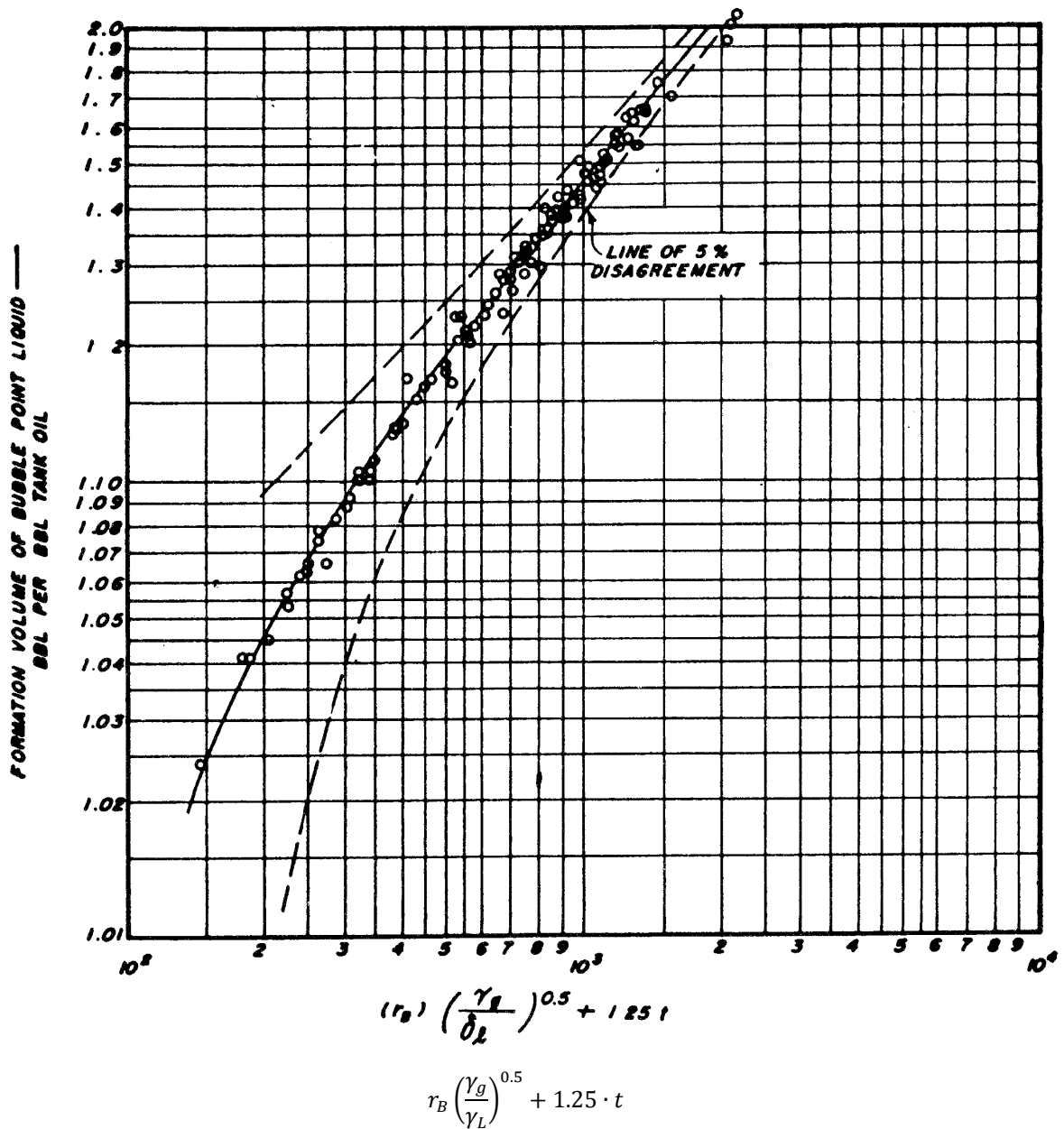


Fig. 9.6. Formation Volume of Bubble Point Liquids from Gas-Oil Ratio, Dissolved Gas Gravity, Tank Oil Specific Gravity, and Temperature.

**Example 9.4.**

To compute the density of a saturated liquid from gas solubility and tank oil gravity data using Correlation of California Crudes.

Given:

	Primary trap	Secondary trap	Tank (STO)	Total
Produced gas-oil ratio, ft <sup>3</sup> /bbl STO	414	90	25	529
Produced gas gravity, air = 1	0.640	0.897	1.54	

Tank oil gravity 27.4°API = 0.891 sp.gr

Bubble point pressure 3280 psia at 218°F

**Solution:**

Basis: One barrel tank oil

Average produced gas gravity,  $\gamma_{g,avg}$  :

$$\gamma_{g,avg} = \frac{414 \cdot 0.640 + 90 \cdot 0.897 + 25 \cdot 1.54}{414 + 90 + 25} = \frac{384.2}{529} = 0.726$$

Weight dissolved gas,  $m_g$ :

$$m_g = \frac{529}{379} \cdot (0.726 \cdot 28.97) = 29.3 \text{ lb}$$

Weight tank oil,  $m_{STO}$ :

$$m_{STO} = 0.891 \cdot 5.615 \cdot 62.4 = 312 \text{ lb}$$

Total weight (gas and Oil),  $m_{tot}$ :

$$m_{tot} = m_g + m_{STO} = 341.3 \text{ lb}$$

$$r_B \left( \frac{\gamma_g}{\gamma_L} \right)^{0.5} + 1.25 \cdot t = 529 \cdot \left( \frac{0.726}{0.891} \right)^{0.5} + 1.25 \cdot 218 = 749.9$$

From Fig.9.6, formation volume factor at bubble point (bp) :

$$(B_o)_{bp} = 1.32 \text{ bbl/bbl STO}$$

Liquid density at 3280 psia and 218 °F,  $(\rho_o)_{bp}$ :

$$(\rho_o)_{bp} = \frac{341.3}{1.32 \cdot 5.615} = 46.0 \text{ lb/ft}^3$$

### 9.1.6. Summary of Principal Points

Four methods have been outlined of calculating the density of a liquid at its bubble point condition of 3280 psia and 218°F. For the single liquid used in the illustration, the following answers were obtained.

	Density, lb/ft <sup>3</sup> 3280 psia, 218°F	% error
Experimental	46.4	
Ideal solution method	48.1	+3.7
Partial volume method	45.3	-2.4
Apparent gas density method	47.5	+2.4
California crude correlations	46.0	+0.9

When a liquid contains only minor amounts of methane and ethane (for example, a primary trap liquid or a tank Oil), "average" values of 20.6 lb/ft<sup>3</sup> for methane (0.33 g/cm<sup>3</sup>) and 29.0 lb/ft<sup>3</sup> for ethane (0.45 g/cm<sup>3</sup>) can be used to represent these components in the ideal solution method. This shortens the

calculation by not requiring the evaluation of propane plus properties and the weight % methane and ethane in the system.

Use an " average " apparent liquid density of 29.9 lb/ft<sup>3</sup> at 14.7 psia and 60 °F for dissolved nitrogen.

The information in Fig. 9.3, Fig. 9.4 and Fig 9.6 can be used to determine the compressibility and thermal expansion of reservoir liquids when direct measurements are lacking. For example, using the method outlined in Example 9.4, a density of 46.0 lb/ft<sup>3</sup> was obtained for the liquid at its bubble point. By applying the data of Fig. 9.3, the bubble point liquid is calculated to have an undersaturated density of 46.62 lb/ft<sup>3</sup> at 5000 psia and 218°F. The average compressibility of the liquid over the indicated pressure interval is:

$$46.31 \left( \frac{\frac{1}{46.62} - \frac{1}{46.0}}{5000 - 3280} \right) = -7.8 \cdot 10^{-6} \text{ volume per volume per psi}$$

## 9.2. Standing Charts

Standing has made empirical correlations using production data from California crude oils. 4 different Standing Charts was developed and are given on the next pages. Based on available production data, the charts could be used to estimate properties of natural hydrocarbon mixtures of Gas and Liquid, formation volume factor ( $B_o$ ), bubble point pressure ( $P_b$ ), formation volume factor at bubble point,  $(B_o)_{bp}$ , and specific gravity ( $\gamma_o$ ) of oil with dissolved gas at  $P$  and  $T$ .

### 9.2.1 Formation volume factor ( $B_o$ )

Formation volume factor ( $B_o$ ) for an oil with dissolved gas could be determined using Chart 1, Fig. 9.7. The production data needed is:

- Gas oil ratio:  $R_s$  (SCF/SBBL)
- Dissolved gas gravity:  $\gamma_g$
- Tank oil gravity:  $\gamma_{STO}$  (°API)
- Reservoir temperature:  $T$  (°F)
- Reservoir pressure;  $P$  (PSIA)

The  $B_o$  factor is reported in BBL/SBBL or  $V/V_{sc}$ .

### 9.2.2 Bubble point pressure, ( $P_b$ )

Standing Chart 2 (Fig. 9.8) could be used to estimate the bubble point pressure ( $P_b$ ) for an oil with dissolved gas. The production data needed are;

- Gas oil ratio:  $R_s$  (SCF/SBBL)
- Dissolved gas gravity:  $\gamma_g$
- Tank oil gravity:  $\gamma_{STO}$  (°API)
- Temperature:  $T$  (°F)

The  $P_b$  is reported in PSIA, Fig. 9.8.

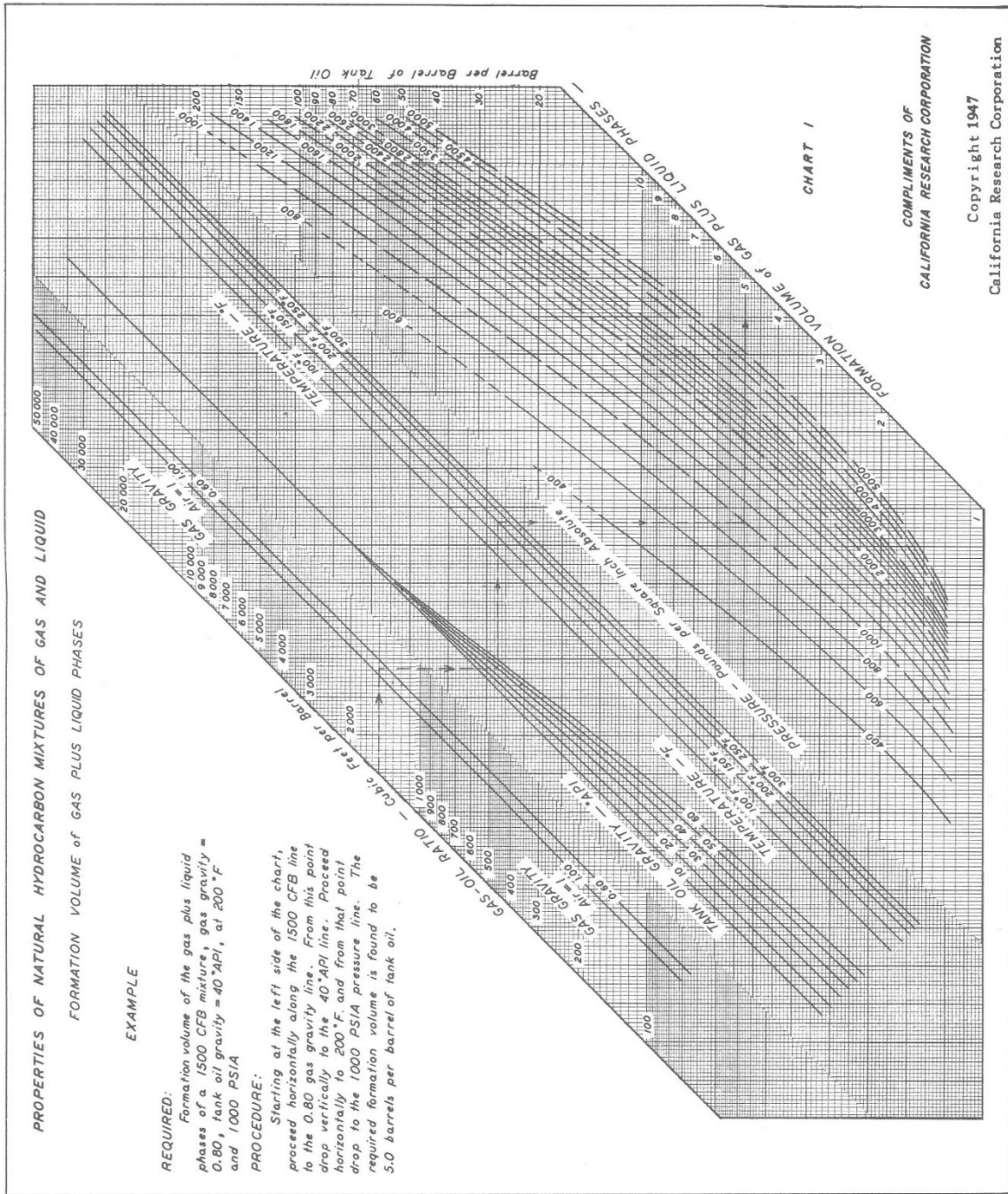


Fig. 9.7. Standing Chart 1. Formation volume factor ( $B_o$ ) for oil with dissolved gas.



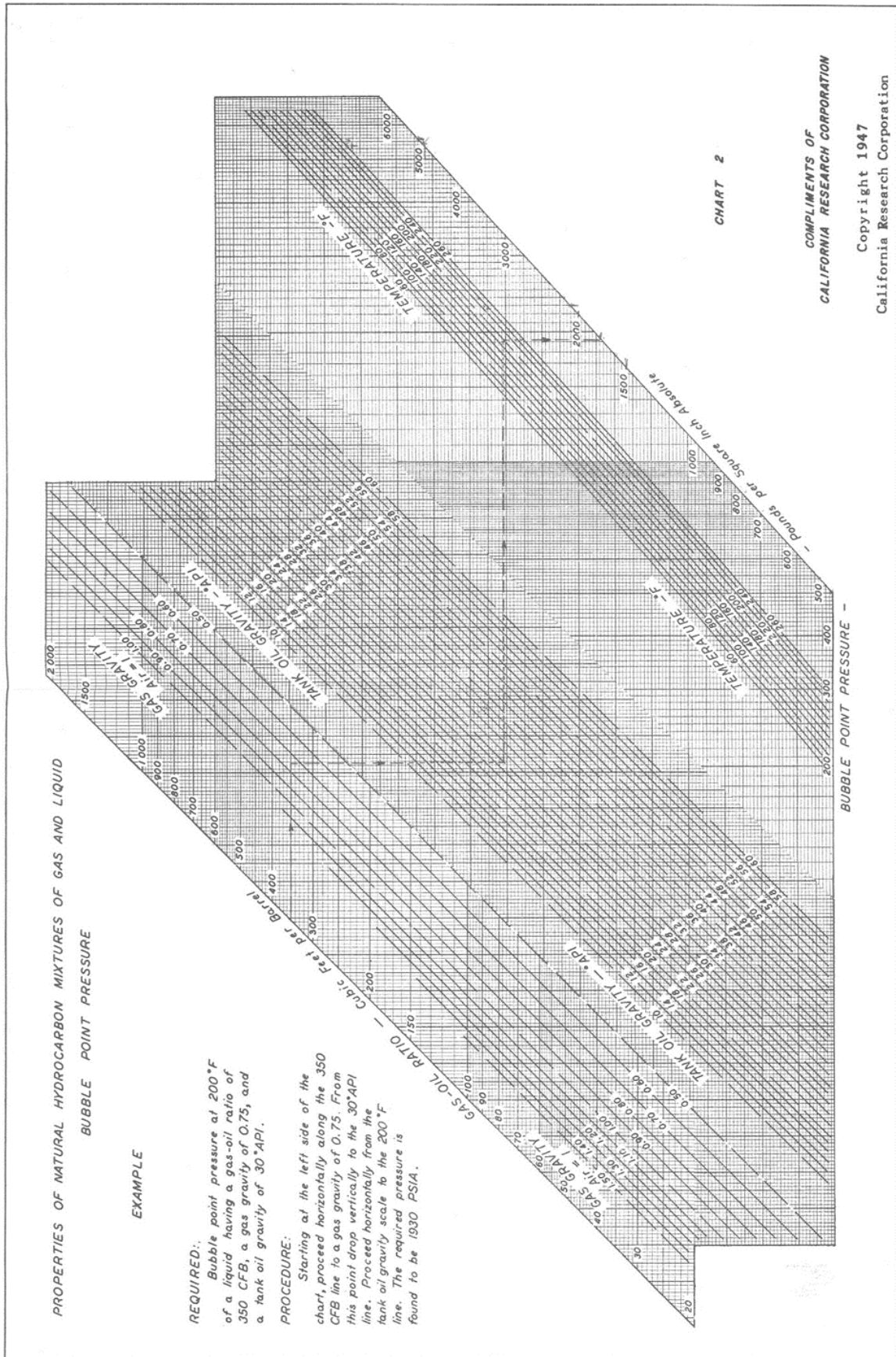


Fig. 9.8. Standing Chart 2. Bubble point pressure for oil with dissolved gas.

### 9.2.3 Formation volume factor at bubble point ( $(B_o)_{bp}$ )

Formation volume factor at bubble point,  $(B_o)_{bp}$ , for an oil with dissolved gas could be determined using Chart 3, Fig. 9.9. The input production data needed is:

- Gas oil ratio:  $R_s$  (SCF/SBBL)
- Dissolved gas gravity:  $\gamma_g$
- Tank oil gravity:  $\gamma_{STO}$  ( $^{\circ}$ API)
- Temperature:  $T$  ( $^{\circ}$ F)

The  $(B_o)_{bp}$  factor is reported in BBL/SBBL or  $V/V_{sc}$ .

A curve fit program for Chart 3 is listed in Standing p. 124;

$$(B_o)_{bp} = 0.9759 + 12 \cdot (10^{-5}) \left\{ R_s \cdot \left( \frac{\gamma_g}{\gamma_{STO}} \right)^{0.5} + 1.25 \cdot T \right\}^{1.2} \quad (9.9)$$

### 9.2.4 Density of a bubble point liquid (oil with dissolved gas) ( $\gamma_o$ )

The specific gravity or density of oil with dissolved gas at bubble point  $(\gamma_o)_{bp}$  can be determined from Standing Chart 4, Fig.9.10. The production data needed are;

- Gas oil ratio:  $R_s$  (SCF/SBBL)
- Dissolved gas gravity:  $\gamma_g$
- Tank oil gravity:  $\gamma_{STO}$  ( $^{\circ}$ API)
- Formation volume factor  $(B_o)_{bp}$  (From Chart 3)

The specific gravity (water = 1) at bubble point  $(\gamma_o)_{bp}$  is reported in lb/ft<sup>3</sup>.

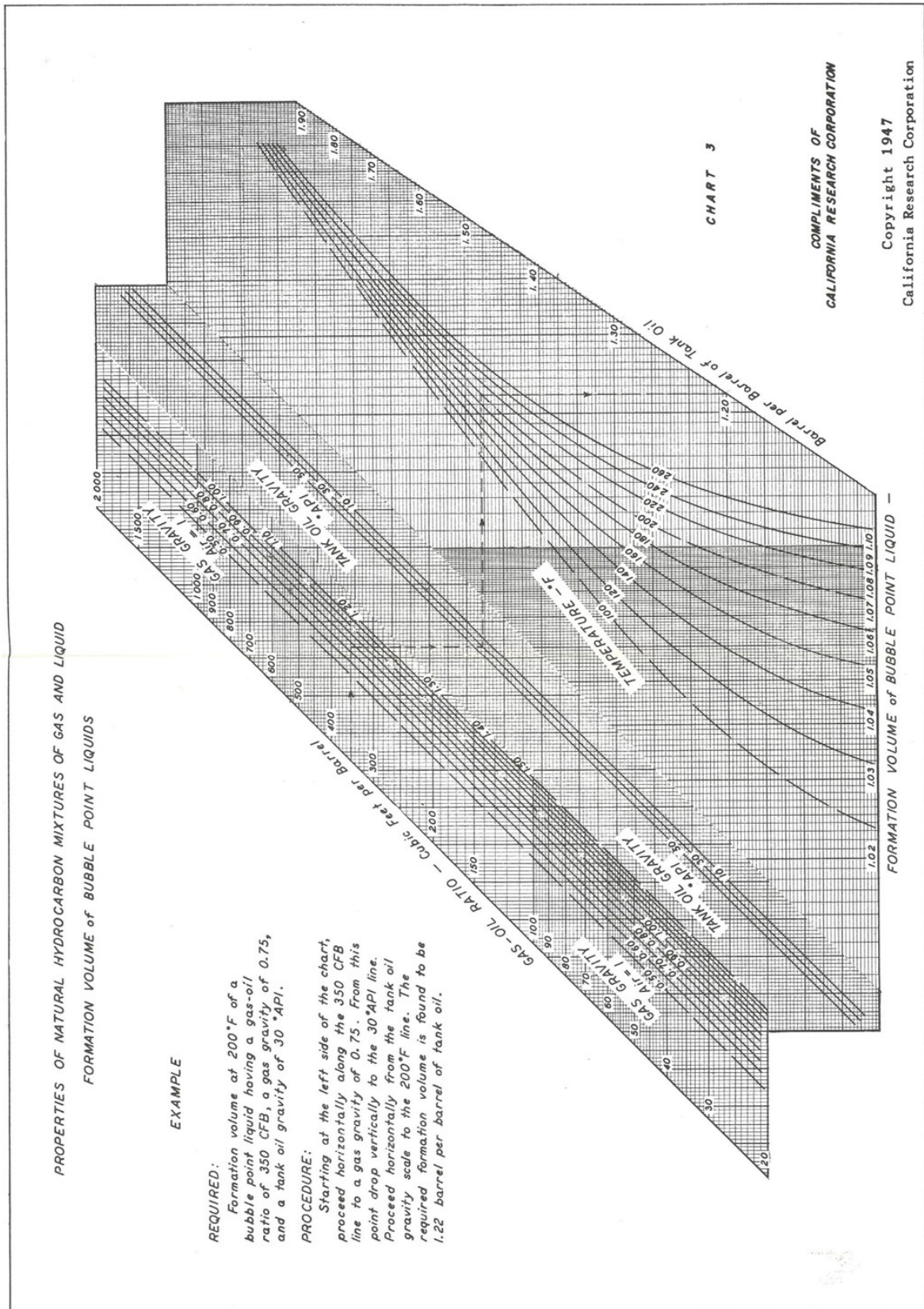


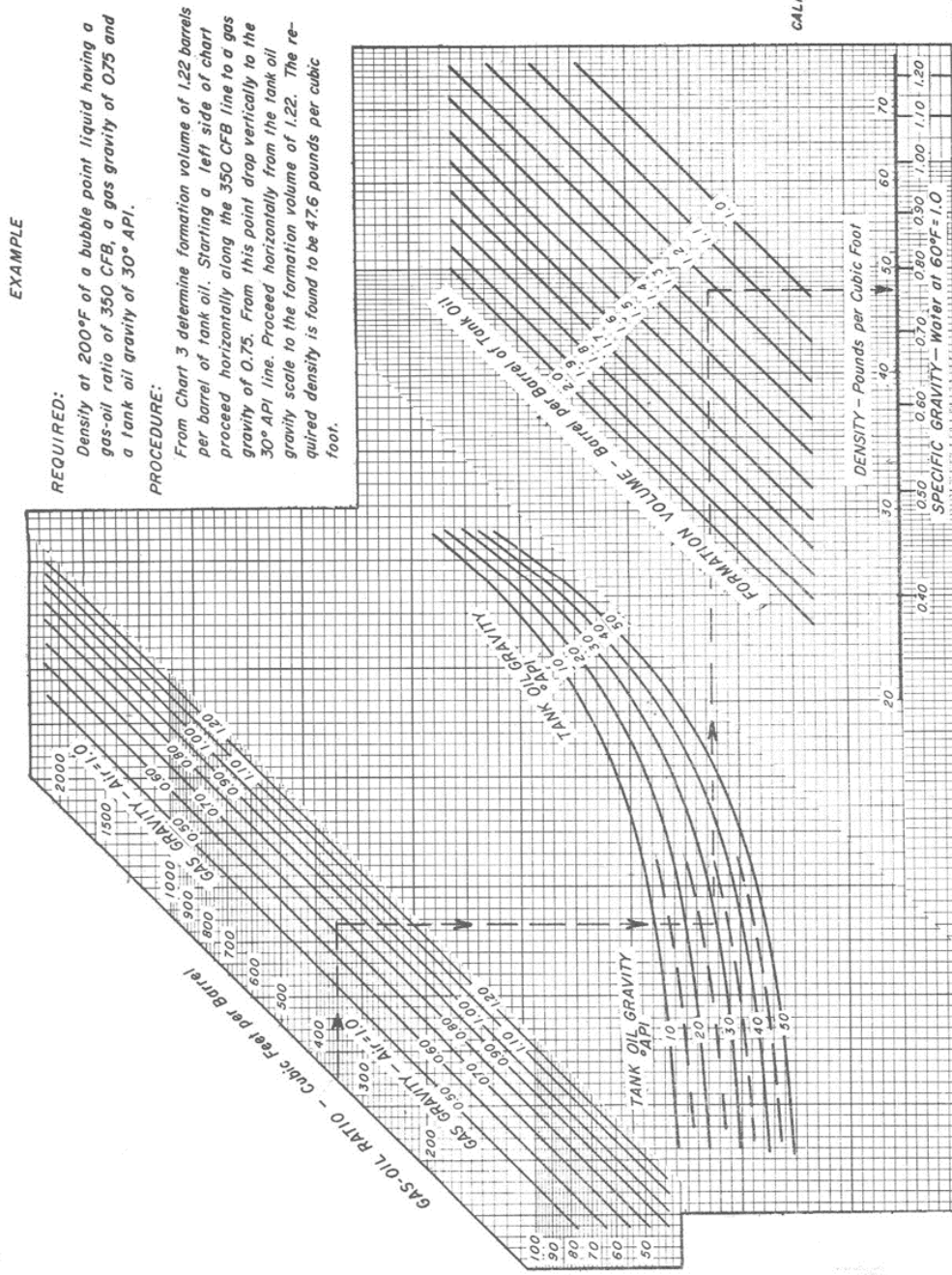
Fig. 9.9. Standing Chart 3. Formation volume factor at bubble point ( $B_o$ )<sub>bp</sub>.

A curve fit program is listed in Standing p. 124;

$$(B_o)_{bp} = 0.9759 + 12 \cdot (10^{-5}) \left\{ R_s \cdot \left( \frac{\gamma_g}{\gamma_{STO}} \right)^{0.5} + 1.25 \cdot T \right\}^{1.2}$$

PROPERTIES OF NATURAL HYDROCARBON MIXTURES OF GAS AND LIQUID

DENSITY and SPECIFIC GRAVITY of MIXTURES



EXAMPLE

REQUIRED:

Density at 200°F of a bubble point liquid having a gas-oil ratio of 350 CFB, a gas gravity of 0.75 and a tank oil gravity of 30° API.

PROCEDURE:

From Chart 3 determine formation volume of 1.22 barrels per barrel of tank oil. Starting a left side of chart proceed horizontally along the 350 CFB line to a gas gravity of 0.75. From this point drop vertically to the 30° API line. Proceed horizontally from the tank oil gravity scale to the formation volume of 1.22. The required density is found to be 47.6 pounds per cubic foot.

CHART 4

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Fig. 9.10. Standing Chart 4. Oil density at bubble point.

## 9.4 In-place reserve calculations

Based on constant mass expansion (CME) and separator test data from the laboratory, the most correct in place reserve calculations could be made for oil with dissolved gas or gas condensates.

CME measurements above saturation pressure ( $P_s$ ) we will give us the one phase fluid volume at saturation point  $(V_o)_s$  and initial reservoir pressure  $(V_o)_i$  which is needed for calculation of volume factors.

Reservoir fluids at saturation pressure ( $P_s$ ) and reservoir temperature ( $T_{res}$ ) are brought to standard conditions in a differential process through separators at specified  $T$  and  $P$ . The test is important for adjusting separator conditions for optimal oil production (tune  $P_{sep}$  to obtain a minimum in  $(GOR)_{tot}$  at given  $T_{sep}$ ). The separator test will also give the most correct volume of produced gas,  $(V_{g(SC)})_d$ , and produced liquid ( $V_{STO}$ ) for in-place reserve calculations, initial oil in place ( $IOIP$ ) and initial gas in place ( $IGIP$ ).

Volume of gas produced in a 3 stage separator system is:

$$(V_{g(SC)})_d = (V_{g(SC)})_1 + (V_{g(SC)})_2 + (V_{g(SC)})_3 \quad (9.10)$$

Gas-oil ratio :

$$(GOR)_{tot} = \frac{(V_{g(SC)})_d}{V_{STO}} \quad (9.11)$$

Oil formation volume factors ( $B_o$ ) at initial pressure ( $P_i$ ) and saturation pressure ( $P_s$ ):

$$(B_o)_i = \frac{(V_o)_i}{V_{STO}} \quad (9.12)$$

$$(B_o)_s = \frac{(V_o)_s}{V_{STO}} \quad (9.13)$$

The initial Oil in place ( $IOIP$ ) and the initial gas in place could then be estimated:

$$(IOIP) = \frac{(HCPV)}{(B_o)_i} \quad (9.14)$$

$$(IGIP) = \frac{(GOR)_{tot}}{(IOIP)} \quad (9.15)$$

With negletable reservoir compaction and inactive aquifer, the hydrocarbon pore volume ( $HCPV$ ) are constant during production, and the production ( $P_{STO}$  and  $P_{g(SC)}$ ) during the pressure decline from  $P_i$  to  $P_s$  could be estimated:

$$P_{STO} = \frac{(HCPV)}{(B_o)_i} - \frac{(HCPV)}{(B_o)_s} \quad (9.16)$$

$$P_{g(SC)} = \frac{(GOR)}{P_{STO}} \quad (9.17)$$

The production could be several % of the  $HCPV$ , dependant of the  $GOR$  and the degree of undersaturation.

Typical production strategies for undersaturated oil fields is to produce the reservoir down to  $P_b$  and maintain the reservoir pressure by water injection to avoid 2-phase fluid flow inside the reservoir. This could allow an initial primary production for months or a few years before the water injection need to start.

## 9.3 Viscosity of oils with dissolved gas

### 9.3.1 Experimental data

In general, the viscosity of live oils ( $\mu_o$ ) is determined experimentally and the data are presented in the PVT-report. The viscosity is determined at reservoir temperature at pressure intervals relevant for oil production. A typical variation of  $\mu_o$  as a function of pressure is illustrated in Fig. 9.11.

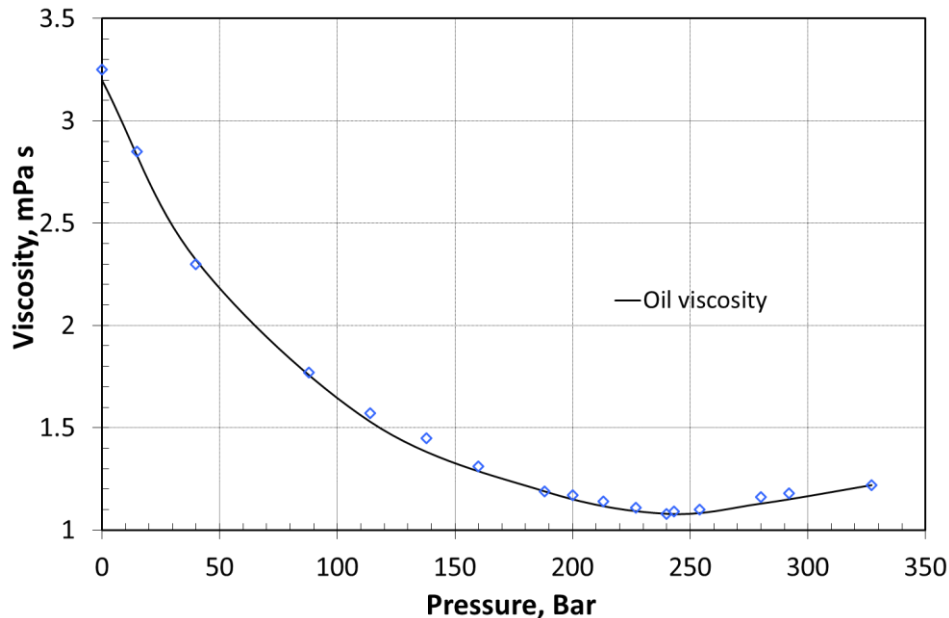


Fig. 9.11. Viscosity of liquid phase ( $\mu_o$ ) for an oil with dissolved gas at reservoir temperature 71.6 °C as a function of reservoir pressure ( $P$ ).

### 9.3.2 Oil Viscosity correlations

If experimental data for the viscosity is not available, there are correlations developed for calculating viscosities of crude oil with dissolved gas.

Several methods have been suggested for estimating the viscosity of hydrocarbon mixtures. In the simulation of processes related to oil and gas production, viscosity correlations are needed that are applicable to a wide range of hydrocarbon mixtures and process conditions.

The viscosity correlation that has the most widespread use in flow models for petroleum mixtures is probably the correlation of Jossi et al. (1962) in the form suggested by Lohrenz et al. (1964). Gas and liquid viscosities are related to a fourth-degree polynomial in the reduced density,  $\rho_r = \frac{\rho}{\rho_c}$ .

The viscosity may alternatively be calculated using a modified form of the corresponding states method. The starting point is the properties at the critical point (e.g., the critical viscosity,  $\eta_c$ , the critical density,  $\rho_c$ , the critical pressure,  $P_c$ , and the critical temperature,  $T_c$ ). Away from the critical point, the properties may be expressed in terms of reduced properties ( $\eta_r$ ,  $\rho_r$ ,  $P_r$ ,  $T_r$ , etc.), which are defined as the properties at the actual conditions, divided by those at the critical point.

By expressing the viscosity as a function of the density, the calculated viscosity becomes very sensitive to the results obtained for the density. Especially for high viscous fluids this may lead to severe errors for the calculated viscosity.

it is possible to predict the viscosity at  $P > P_b$  using the method of Thodos.

Three different charts are used:

- a. The viscosity of STO is determined at reservoir temperature as a function of  $\gamma_{STO}$ .
- b. The viscosity of gas-saturated oil at  $T_{res}$  and  $P_b$  is then determined.
- c. Finally, the viscosity increase from  $P_b$  to the under-saturated pressure is determined.

### 9.3.3 Viscosity of stabilized oils - Kinematic viscosity

The viscosity of lubricants and stabilized oils without dissolved gas could be estimated by the use of a kinematic viscosity approach.

The dynamic viscosity in centipoises,  $\mu$ , divided by the density,  $\rho$ , in  $\text{g/cm}^3$  is defined as the kinematic viscosity,  $\nu$ , in centistokes (cSt):

$$\nu = \frac{\mu}{\rho} \quad (9.18)$$

The kinematic viscosity exhibits a linear trend when plotted as a function of temperature on a special chart available from the American Society of Testing Materials (ASTM). Thus, observations at only two temperatures are enough to determine the viscosity of stabilized oils at other temperatures.

## 9.4 References

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