

11. RESERVOIR FLUID CALCULATIONS - GAS-CONDENSATE SYSTEMS

A gas-condensate system will exhibit isothermal retrograde behavior in the temperature range that is of interest to petroleum engineering studies. Such a system has a critical temperature lower than the reservoir temperature or the surface temperature at which it is separated into gas and liquid. Production of a condensate system through surface traps usually results in gas-oil ratios greater than 5 000 St.ft³/Sbbl* and tank oil gravity between 45 and 70° API. The tank oils usually are white or only slightly colored. A definition of the upper value of gas-oil ratio of a gas-condensate system is arbitrary; usually a system having a ratio greater than 100 000 St.ft³/Sbbl (0.43 gal/MCF) is considered a dry gas.

*The value 5 000 St.ft³/Sbbl assumes production of only vapor from the reservoir. The production of both vapor and liquid from a reservoir to give 5 000 St.ft³/Sbbl (or more) does not necessarily indicate a condensate system. Pressure decline in gas-condensate reservoirs results in formation of a liquid phase within the reservoir rock pores as seen in Fig. 11.1 .

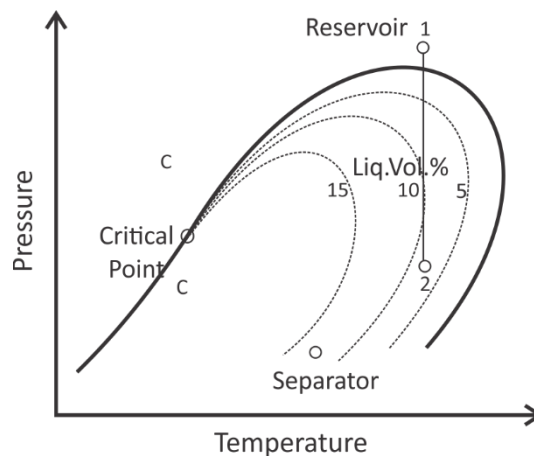


Fig. 11.1 . Phase diagram of gas condensate.

The operation of processing production from gas-condensate reservoirs for the heavier hydrocarbon components and returning the light components to the reservoir (cycling) has become increasingly popular. It often is stated that the purpose of cycling is to maintain the reservoir pressure near its original pressure in order to remove the heavier hydrocarbon components which otherwise would form the retrograde liquid.

The PVT behavior of gas-condensate systems that are representative of reservoir fluids can be determined easily by laboratory tests. Information that usually is determined in the laboratory includes compressibility factors of the reservoir vapor phase and surface gases, quantity of liquid formed at reservoir and atmospheric conditions as a result of retrograde action, and data from which the decline in reservoir pressure resulting from production of reservoir vapor can be calculated directly. Less is known of the composition of reservoir vapor and retrograde liquid during pressure decline than of their volumetric behavior. Laboratory experiments to determine the composition of the two phases are laborious and time-consuming and have received little attention. Calculated compositions and phase volumes based solely on equilibrium constant data can be greatly in error. Probably the best estimate of phase composition can be obtained by combining laboratory-determined volumetric data with published equilibrium constant data. The phase composition so computed will not be greatly in error and will indicate the trend that the reservoir phases will follow.

11.1 Laboratory Tests

The laboratory apparatus used in condensate studies differs from the apparatus used in dissolved gas studies for two reasons:

- (1) The dew point pressure of most condensate systems cannot be detected by a sharp change in the pressure-volume relation of the system. (Bubble point pressure is determined most easily by this means.)
- (2) The liquid phase constitutes only a small part of the total volume in the laboratory testing equipment. It is necessary therefore to have methods of precisely measuring small liquid quantities.

One of the most satisfactory PVT cells is one with glass windows to permit the visual observation of dew points and the quantity of retrograde liquid at pressures less than the dew point pressures. The type of cell in which the liquid phase volume is determined by locating the interface between vapor and liquid by thermal methods* is equally satisfactory when liquid volumes greater than two cubic centimeters are involved. However, the second type of cell is inferior to a glass cell when an accurate measure is desired of a system's dew point pressure, because the dew point, pressure must be obtained extrapolation of a liquid phase volume pressure curve to zero liquid volume.

* This type cell and its operation are described in more detail in Chapter 7 and by Sage and Lacey.[1]

11.1.1 Flash Tests.

Flash tests are carried out on gas condensate systems to determine volumetric behavior of the systems under surface trap conditions, under well tubing conditions, and under reservoir conditions. As previously defined, a flash test signifies constant system composition during the progress of the test. The tests are made most often on systems prepared from surface samples of trap fluids recombined in their produced ratios.

It is not difficult to measure phase volumes in a laboratory PVT cell at the higher pressures corresponding to well tubing pressures and to reservoir pressures. The weight of hydrocarbon material in the cell during the tests can be determined accurately at the completion of the tests by measuring the quantity and gravity of the gas and liquid removed from the cell. Often the gas and liquid are removed from the cell under the pressure and temperature conditions existing in the field traps. When this is done, the volumetric data can be given in terms of formation volumes, or in terms of volumes under reservoir conditions per MCF of surface trap gases. For many purposes a specific volume or density value obtained from the measured weight of the system and its measured volume at specified pressures and temperatures is more valuable in engineering calculations than the formation volume or "reservoir" volume.

It is not always feasible to determine experimentally the effect of surface trap pressures and temperatures on the properties and volumes of surface products. Size limitations of the usual laboratory PVT equipment often make it impossible to expand high gas-oil ratio systems to pressures of the order of 250 psia and still have sufficient liquid phase in the cell to measure accurately its volume. Therefore, in many condensate systems, it is necessary to use equilibrium calculation methods to evaluate the effect of surface trapping conditions on the quantity and gravity of tank oil and trap gases. Because of the inaccuracies of the calculation methods, the absolute values obtained by this process may not agree with field measurements, although the trend shown by the calculation usually is correct.

11.1.2 Differential Tests.

By definition, a differential process is one in which the over-all composition of the system varies as the process proceeds. The differential tests made, on gas-condensate systems are based on the assumption that the retrograde liquid formed in the reservoir rock is immobile. In line with this assumption, the laboratory test procedure usually is as follows:

- (1) A system having a composition as close as possible to that of the original reservoir fluid is constructed in the laboratory PVT cell. The system is brought to the original reservoir pressure and temperature or to the system's dew point pressure at the reservoir temperature.
- (2) Starting at the dew point or reservoir pressure, equilibrium vapor phase material is removed from the top of the cell. At several points during the process, measurements are made of the cell pressure, the volume of retrograde liquid formed in the cell, and the cumulative volume of vapor removed from the cell. This step simulates the production of vapor phase from the reservoir rock.
- (3) During the vapor removal process, the total volume occupied by the system is kept constant. This is equivalent to assuming that the reservoir volume does not change during its producing life because of water encroachment. The assumption, of course, may be invalid for many reservoirs, but its invalidity does not detract from the usefulness of the laboratory data because adjustments can be made to the laboratory data where the true reservoir condition is known.

The fundamental assumption of the differential tests concerning immobility of retrograde reservoir liquid is believed reasonable in view of the facts that most condensate systems reported in the literature yield less than 10 % retrograde liquid by volume during pressure decline at reservoir temperatures, and that the effective permeability of reservoir rocks to this phase usually is zero in the saturation range of interest.

There is good reason to believe, however, that the above assumption is not valid in the region of the well bore. A higher liquid saturation probably exists near the well bore than out in the reservoir because of the larger pressure gradient and the increased fluid flux density near the well bore. Over a period of time the liquid saturation near the well bore may increase to the point where the pressure gradient may overcome the capillary forces retaining the liquid in the rock. The liquid phase will then be produced in the well concurrently with vapor material. This local action in the region of the well bore is of interest primarily to the problem of recombining surface samples to obtain a system that is representative of the system in the major portion of the reservoir.

11.2 Condensate System Behavior in the Single-Phase Region

The volumetric behavior of a condensate system under pressure and temperature conditions that permit only a single phase is similar to the behavior of dry gases previously discussed in chapter 8. The volume of the system can be calculated by use of the gas law equation or by any one of several empirical correlations. The sections to follow will outline methods of evaluating compressibility factors for condensate system calculations, as well as present some empirical correlations.

11.2.1 Compressibility Factors.

The volume occupied by a condensate system at pressures and temperatures outside the phase diagram can be calculated from the equation $PV = ZnRT$. The compressibility factor, Z , can be determined from the pseudo reduced pressure and temperature correlations shown in Fig. 8.4, provided that proper pseudo constants are used to represent the relatively large proportion of high molecular weight components present in the system.

Standing and Katz [2] have correlated pseudo-critical values and the molecular weight of the heptanes plus fraction of twelve high pressure equilibrium vapors. The pseudo-critical pressures were found to be roughly 50 psi higher than true critical pressures of equivalent molecular weight normal paraffin hydrocarbon. The pseudocritical temperatures were found to be about 30°F higher than the true critical temperatures of equivalent molecular weight paraffin hydrocarbons. Mathews, Roland, and Katz [3] found that volumetric calculations could be made with increased accuracy if the specific gravity, as well as the molecular weight, were used in evaluating the pseudo-critical constants of heptanes plus fractions. The correlation presented by these authors is reproduced in Fig. 11.2 and is recommended for use where both the specific gravity and molecular weight of the heptanes plus fraction (C7+) are known.

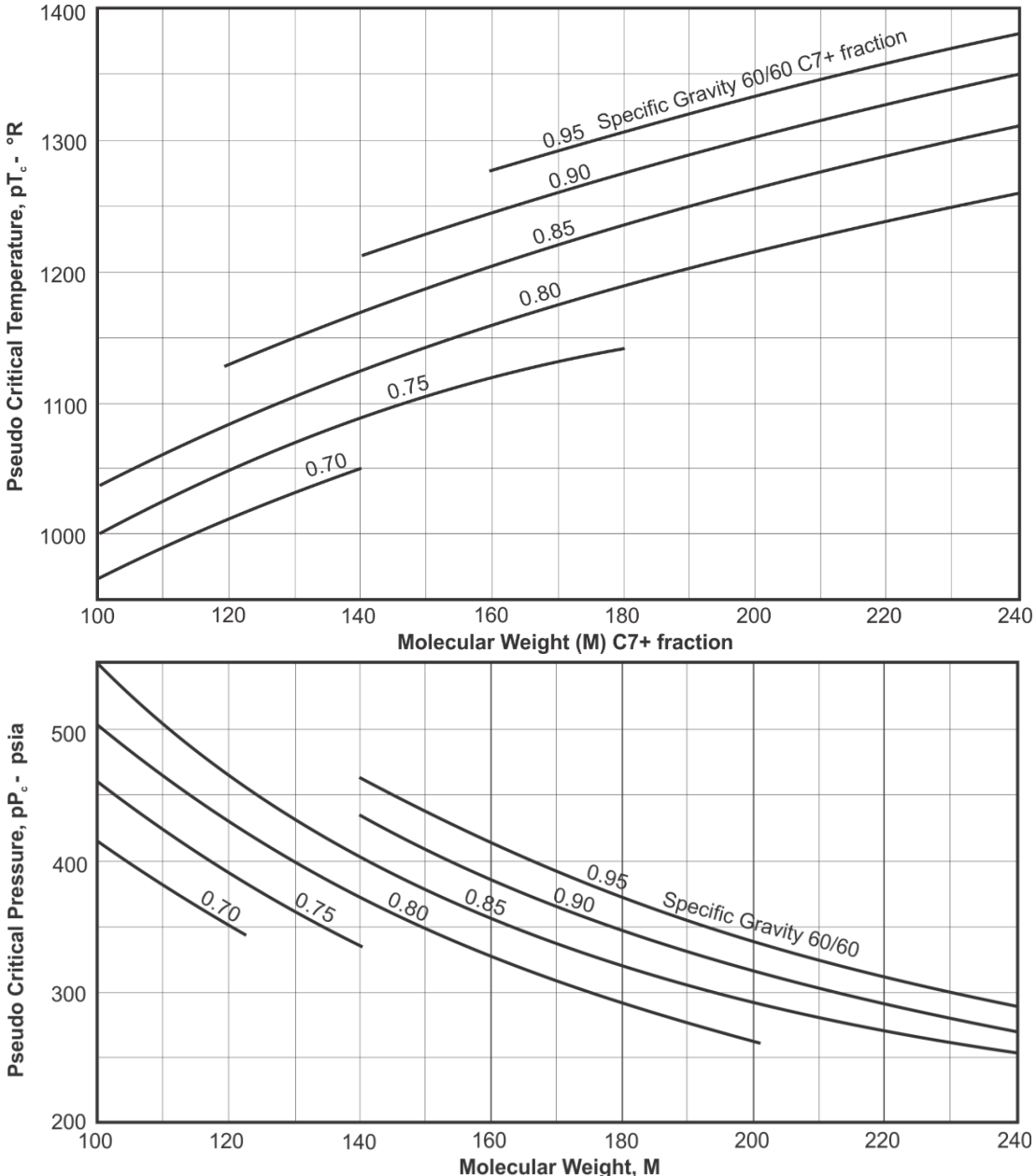


Fig. 11.2. Pseudo-critical Temperatures (pT_c) and Pressures (pP_c) for C7+ fraction (Heptanes and Heavier) when the specific gravity (γ_{C7+}) and Molecule weight of C7+ fraction (M_{C7+}) is known. [3]

Where neither specific gravity nor molecular weight is known, it is necessary to assume pseudocritical values. A reasonable assumption is that the molecular weight of the heptanes plus is equivalent to that

of nonane, which is 128. By adding the 50 psi and 30°F found by Standing and Katz to nonane's true criticals, the pseudo-critical values are 385 psia and 1100°F.

The compressibility factor, Z, as normally used in the Eq. 11.1;

$$PV = ZnRT \quad (11.1)$$

or

$$\bar{V} = \frac{ZRT}{PM} \quad (11.2)$$

should be applied only in the pressure and temperature range in which the system remains in a single phase. If the equation is used at pressures and temperatures where two phases exist, \bar{V} in the Eq. 11.2 expresses the average specific volume of the two-phase system. In many instances, the average specific volume of the system when in two phases is close to the specific volume calculated as though the system existed as a single phase. This is particularly true in the pressure interval 2000 to 6000 psia when the liquid phase represents less than 5% of the total system volume.

11.2.2 Pseudo-critical Constant-Gravity Relations of Condensate Systems.

The basis for correlations of the pseudo-critical pressure and temperature with molecular weight of a gas has been explained previously.

The correlations found by Standing and Katz [2] for a group of condensate vapors was presented together with gasses, and could be used in calculations pertaining to the volumetric properties of condensate systems

In an effort to simplify condensate system calculations, Rzasa and Katz [4] developed a chart that shows a relationship between the ratio of well fluid gravity (as a vapor) to trap gas gravity and the barrels of surface condensate per million cubic feet of surface gas. Such a chart has considerable utility, as one often knows the surface trap gas gravities, but for calculation purposes requires the gravity of the system in the well or in the reservoir rock. Rzasa and Katz' correlation contained only a single curve, but theoretically such a correlation also should contain variables of trap gas gravity, condensate liquid gravity, and condensate molecular weight. These variables are related by the equation:

$$\frac{\text{well fluid gravity}}{\text{trap gas gravity}} = \frac{M_L}{28.97 \cdot \gamma_g} \left(\frac{76.4 \cdot S \cdot \gamma_g + 350 \cdot \delta_L}{2.64 \cdot S \cdot M_L + 350 \cdot \delta_L} \right) \quad (11.3)$$

where:

M_L = molecular weight of condensate (tank oil)

δ_L = specific gravity of condensate (tank oil)

γ_g = gravity of trap gas (air = 1)

S = MCF surface gas per barrel condensate

Fortunately, there exists a fair correlation of molecular weight with the specific gravity of natural condensates so that Eq. 11.3 can be simplified to contain only data which can be determined in the field. The result of such simplifications is the calculating chart presented in Fig. 11.3. The rough correlation of condensate molecular weight and gravity used also is shown in Fig. 11.3.

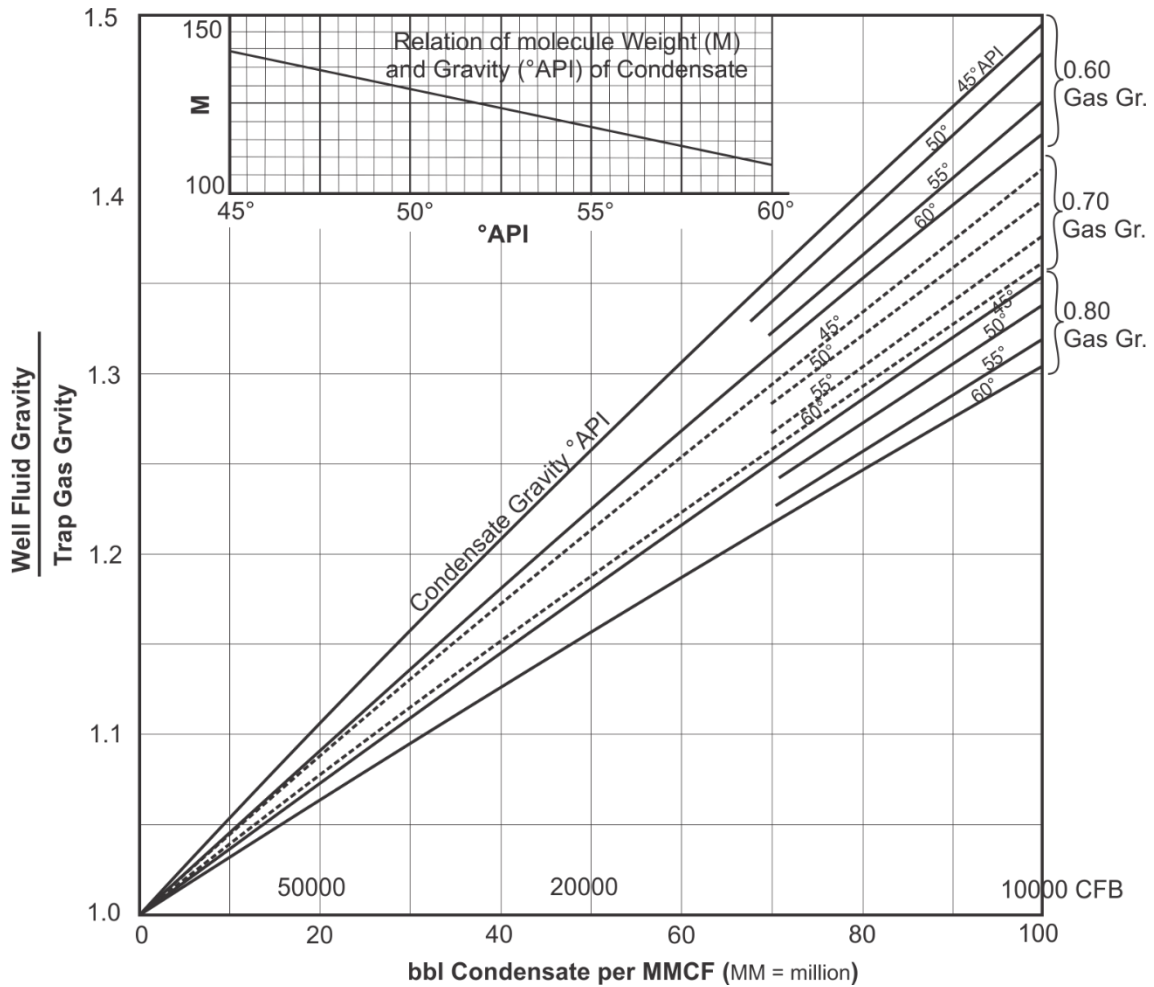


Fig. 11.3. Effect of Condensate Volume on the Ratio of Surface Gas Gravity to Well Fluid Gravity

An example of using the data presented in Fig. 11.3 is given in Example 11.1 .

A. L. Vitter, Jr.[5] has suggested an equation similar to Eq. 11.3 by which to correct the observed trap gas gravity to a well fluid gravity. By assuming that all condensates (tank oils) have an equivalent vapor volume of 200 ft³ gas/ft³ condensate, he obtained the following equation:

$$\gamma_w = \frac{\gamma_g + \frac{4591 \cdot \delta_L}{1000 \cdot S}}{1 + \frac{1123}{1000 \cdot S}} \quad (11.4)$$

Where:

γ_w = gravity of well fluid (air = 1)

Well fluid gravities calculated by this equation do not differ materially from those calculated by use of Eq. 11.4 or Fig. 11.3. In the example given in Example 11.1, a well fluid gravity of 0.896 was obtained by Vitter's equation compared to 0.917 obtained by use of Fig. 11.3. Better comparisons are obtained at high gas-oil ratios and high °API condensate gravities.

Example 11.1 .

Estimation of Gas-in-place at Elevated Pressure and Temperature using Well Gravity to Evaluate Pseudo-critical Constants.

Estimate the standard cubic feet (SCF) of gas in place in a reservoir that has the following properties:

Reservoir pressure	3 000 psia
Reservoir temperature	240°F
Rock porosity	30%
Interstitial water	27 %
Tank oil production	323 bbl/Day
Tank oil gravity	45°API
Primary trap gas rate	3 765 MCF/Day
Primary trap gas gravity	0.65 (air = 1)
Tank vapor rate	169.2 MCF/Day
Tank vapor gravity	1.25 (air = 1)

Solution:

Basis: One acre-foot sand volume

Average separator gas gravity

$$(\gamma_g)_{avg} = \frac{3765 \cdot 0.65 + 169 \cdot 1.25}{3765 + 169} = 0.677$$

bbl condensate per MMCF gas:

$$\frac{V_{STO}}{V_g} = \frac{323}{3.934} = 82.1 \text{ bbl/MMCF}$$

From Fig. 11.2;

$$\frac{\text{well fluid gravity}}{\text{trap gas gravity}} = 1.35$$

Well fluid gravity

$$= 1.35 \cdot 0.677 = 0.917$$

From Fig. 8.6 in Gas chapter;

$${}_pT_c = 430^\circ\text{R}, \quad {}_pP_c = 650 \text{ psia},$$

giving

$${}_pT_r = \frac{460 + 240}{430} = 1.63, \quad {}_pP_r = \frac{3000}{650} = 4.62$$

Z are found from Fig. 8.4 in the gas chapter; giving:

$$Z = 0.840$$

The reservoir HCPV per acre-foot of rock is;

$$43560 \cdot 0.30(1 - 0.27) = 9540 \text{ ft}^3$$

Total moles HC per acre-foot at reservoir conditions

$$n = \frac{PV}{ZRT} = \frac{3000 \cdot 9540}{0.840 \cdot 10.73 \cdot 700} = 4540 \text{ lb moles}$$

Now, if all this quantity of HC were produced as gas at the surface, it would represent:

$$4540 \cdot 379 = 1730 \text{ MCF per acre - foot}$$

However, as a portion will be produced as tank oil, it is necessary to make allowance for this fact.

One barrel of 45°API tank oil ($\delta_{STO}=0.802$) having a molecular weight of 140 would correspond to:

$$n_{STO} = \frac{m_{STO}}{M_{STO}} = \frac{5.615 \cdot 62.37 \cdot 0.802}{140} = 2.01 \text{ lb moles}$$

At the present producing ratio of 12 170 ft³/bbl, the total surface gas amounts to;

$$n_g = \frac{V_g}{V_m} = \frac{12\ 170}{379} = 32.1 \text{ lb moles}$$

Therefore, under the present producing of the reservoir fluid the mole fraction of surface gas (V) is :

$$V = \frac{32.1}{32.1 + 2.01} = 0.941$$

and mole fraction of STO (L):

$$L = 0.059$$

Surface gas :

$$(V_g)_{sc} = 0.941 \cdot 4540 \cdot 379 = 1620 \text{ MCF/acre - foot}$$

Surface tank oil (STO)

$$V_{sto} = \frac{0.059 \cdot 4540}{2.01} = 133 \text{ bbl/acre - foot}$$

11.2.3 Empirical Behavior in the Single-Phase Region.

Sage and Olds correlation

Sage and Olds [6] correlated the volumetric behavior of five condensate systems produced from California fields, the producing gas-oil ratios of the systems were in excess of 5 000 ft³/bbl, and the tank oil gravities were between 52° and 63°API. An empirical relation was determined which expressed the formation volume of the system as a function of pressure, temperature, and gas-oil ratio. The relation is:

$$u = A \frac{r \cdot T}{P} \quad (11.5)$$

where

P = pressure (psia)

T = temperature (°R)

r = gas oil ratio, ft³/bbl STO

u = formation volume (ft³/bbl STO)

A = empirical coefficient, as tabulated in Tbl. 11.1 .

Tbl. 11.1 Coefficient A used to Calculate Formation Volume of Gas-condensate Systems, Method by Sage and Olds [6] for use with Eq. 11.5.

1	2	3	4	5	6	7
Pressure psia	$A \cdot 10^3$					
	100°F	130°F	160°F	190°F	220°F	250°F
800	4.46	4.57	4.67	4.76	4.83	4.87
1000	4.35	4.47	4.59	4.69	4.77	4.82
1250	4.21	4.85	4.49	4.62	4.71	4.77
1500	4.09	4.25	4.41	4.55	4.66	4.73
1750	3.99	4.17	4.34	4.50	4.62	4.71
2000	3.93	4.11	4.29	4.46	4.59	4.69
2250	3.89	4.08	4.26	4.43	4.57	4.68
2500	3.88	4.06	4.25	4.42	4.57	4.68
2750	3.89	4.07	4.26	4.44	4.58	4.69
3000	3.92	4.10	4.29	4.47	4.61	4.71

Theoretically, A is not a function of pressure and temperature only, as indicated by Tbl. 11.1. The gas-oil ratio, the gravity of the gas, oil gravity, and molecular weight are factors which cause minor variation in the value of A . Notwithstanding the apparent neglect of the above factors, the method gives formation value factors that follow the volumetric behavior of the California systems with errors no larger than those obtained when other methods are used. An example of use of this method is shown in Example 11.2.

Example 11.2.

Estimation of Gas-in-place (GIP) at elevated Pressure and Temperature using Method of Sage and Olds. [6]

To estimate the standard cubic feet of gas in a reservoir having the following properties:

Reservoir pressure	3000 psia
Reservoir temperature	240°F
Porosity	30%
Interstitial water	27%
Tank oil production	323 bbl STO/Day
Primary trap gas rate	3765 MSCF/Day
Tank vapor rate	169.2 MSCF/Day

Solution:

(Basis: One acre-foot sand volume)

Producing gas-oil ratio (GOR) :

$$GOR = \frac{(3765 + 169)(10^3)}{323} = 12\,170 \text{ SCF/bbl STO}$$

At 3000 psia and 240°F, (Tbl. 11.1)

$$A = 4.68 \cdot 10^{-3}$$

Formation volume factor (u) (Eq. 11.5 by Sage and Olds):

$$u = A \frac{rT}{P} = \frac{4.68 \cdot 10^{-3} \cdot 12170 \cdot (460 + 240)}{3000} = 13.30 \text{ ft}^3/\text{bbl STO}$$

Reservoir HCPV occupied by one day's production;

$$13.30 \cdot 323 \cdot 5.615 = 24100 \text{ ft}^3/\text{Day}$$

One acre-foot reservoir rock have a HCPV of;

$$HCPV = 43560 \cdot 0.30 \cdot (1 - 0.27) = 9540 \text{ ft}^3/\text{acre} - \text{foot}$$

Therefore, one acre-foot sand contains;

$$\frac{9540}{24100} = 0.396 \text{ of a day's production}$$

This is equivalent to

$$13.30 \cdot (3765 + 169) = 1557 \text{ MSCF surface gas}$$

and;

$$0.396 \cdot 323 = 128 \text{ bbl STO}$$

Standing correlation

Standing [7] has made an empirical correlation of the gas plus liquid phases formation volume factor as shown in Fig. 11.4, based on a calculated factor from production data; Gas-Oil Ratio (r), Gas Gravity (γ_g), Tank Oil Gravity (δ_L), Temperature in °F (t) as expressed in Eq. 11.6;

$$r \frac{(t)^{0.5}}{(\gamma_g)^{0.3}} (\delta_L)^{2.9(10)^{-0.00027r}} \quad (11.6)$$

and Pressure in Psia.

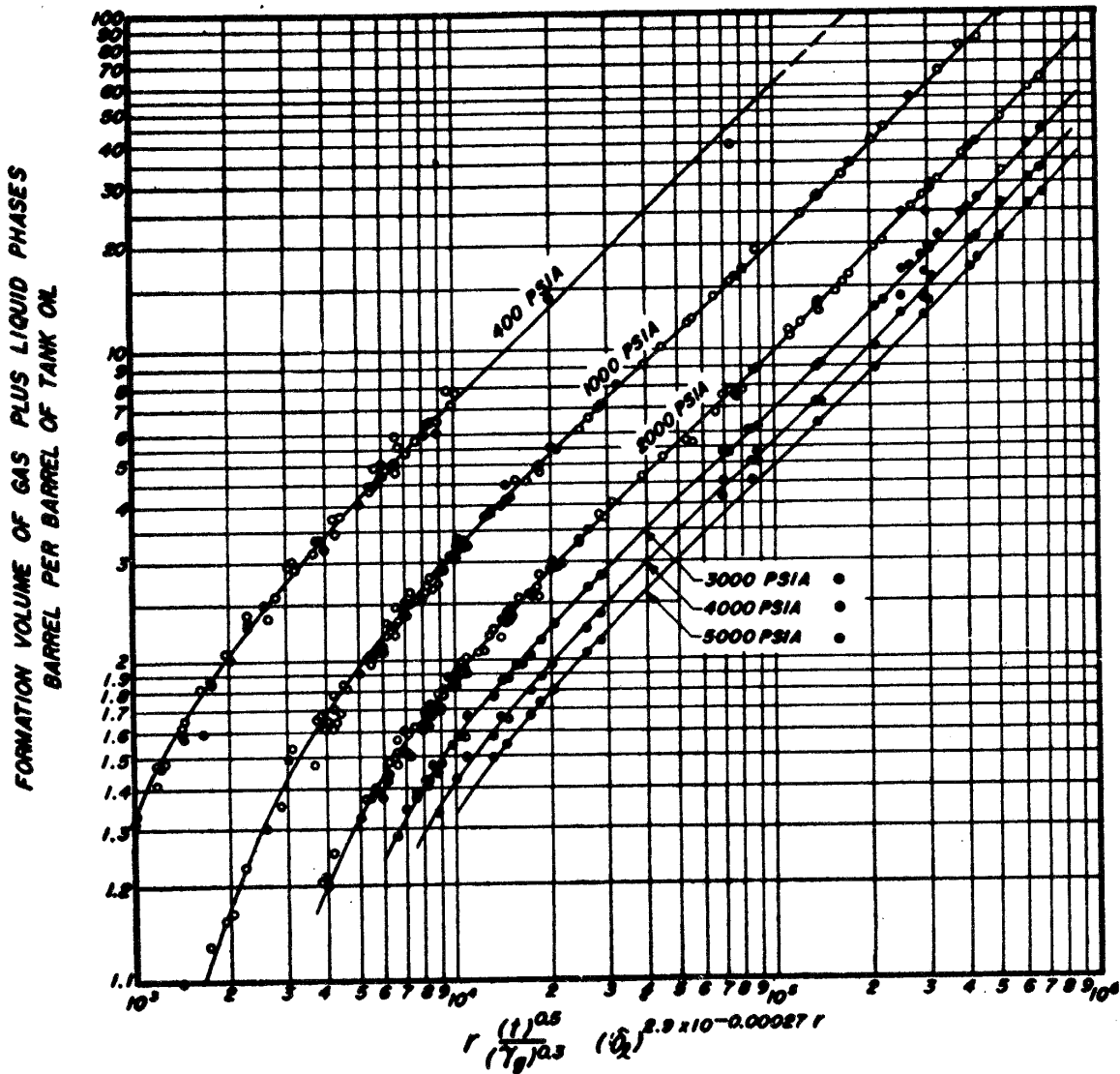


Fig. 11.4. Formation Volume of Gas Plus Liquid Phases from Gas-Oil Ratio, Gas Gravity, Tank Oil Gravity, Temperature, and Pressure

It can be seen that at high gas-oil ratios the effect of the tank oil gravity becomes negligible and the compressibility factor of the system then becomes a function of the trap gas gravity.

The formation volume factors calculated by use of this relation usually are in error less than 3 %. The advantage of the correlation is that it can be used easily in the form of a calculating chart. The formation volume calculation in Example 11.3 is carried out by use of the calculation chart.

Example 11.3.

Estimation of Gas-in-place at Elevated Pressure and Temperature Using Standing's Correlation [7].
 Estimate the standard cubic feet of gas in place in a reservoir having the following properties:

Reservoir pressure	3000 psia
Reservoir temperature	240°F
Rock porosity	30%
Interstitial water	27%
Tank oil production	323 bbl/Day
Tank oil gravity	45 °API
Primary trap gas rate	3765 MCF/Day
Primary trap gas gravity	0.65 (air = 1)
Tank vapor rate	169.2 MCF/Day
Tank vapor gravity	1.25 (air = 1)

Solution:

(Basis: One acre-foot sand volume)

Average separator gas gravity

$$(\gamma_g)_{avg} = \frac{3765 \cdot 0.65 + 169 \cdot 1.25}{3765 + 169} = 0.677$$

Producing gas-oil ratio, GOR;

$$GOR = \frac{(3765 + 169)(10^3)}{323} = 12\,170 \text{ ft}^3/\text{bbl}$$

From Standing correlation (Fig.11.4):

$$r \frac{(t)^{0.5}}{(\gamma_g)^{0.3}} (\delta_L)^{2.9(10)^{-0.00027 \cdot r}} = 12170 \frac{(240)^{0.5}}{(0.677)^{0.3}} (0.8017)^{2.9(10)^{-0.00027 \cdot 12170}}$$

$$= 12170 \frac{15.5}{0.889} 0.999 = 2.125 \cdot 10^5$$

From Fig. 11.4, formation volume at 3000 psia;

$$B_o = 13.5 \text{ bbl/bbl STO}$$

HCPV of one acre-foot reservoir rock;

$$HCPV = 43560 \cdot 0.30 \cdot (1 - 0.27) \frac{1}{5.615} = 1700 \text{ bbl}$$

Therefore, tank oil volume:

$$V_{STO} = \frac{1700}{13.5} = 127 \text{ bbl/acre - foot}$$

surface gas volume:

$$(V_g)_{sc} = 127 \cdot 12170 = 1534 \text{ MCF/acre - foot}$$

11.3 Condensate System Behavior in the Two-Phase Region

11.3.1 Dew Point Behavior.

The dew point, like the bubble point, defines the state at which a system changes from one phase to two phases. The dew point pressure can be calculated, provided the proper equilibrium constants and the composition of the system are known. While it is not too difficult to determine the composition of a system, the blind use of published equilibrium constants can result in errors as great as several thousand pounds per square inch (psi) in the calculated dew point pressures.

Sage and Olds [6] have presented a relation between tank oil gravity, gas-oil ratio, temperature, and dew point pressure for the five San Joaquin Valley, California, condensate systems previously mentioned. Their data, presented in Tbl. 11.2, represent the only published attempt of a general correlation of dew point pressures.

Tbl. 11.2. Relation of Dew Point Pressure of California Condensate Systems.
Data of Sage and Olds [6]

1	2	3	4	5	6	7
Tank Oil Gravity	Gas—Oil Ratio, SCF/bbl					
°API	15000	20000	25000	30000	35000	40000
100°F						
52	4440	4140	3880	3680	3530	3420
54	4190	3920	3710	3540	3410	3310
56	3970	3730	3540	3390	3280	3180
58	3720	3540	3380	3250	3140	3060
60	3460	3340	3220	3100	3010	2930
62	3290	3190	3070	2970	2880	2800
64	3080	3010	2920	2840	2770	2700
160°F						
52	4760	4530	4270	4060	3890	3650
54	4400	4170	3950	3760	3610	3490
56	4090	3890	3690	3520	3380	3270
58	3840	3650	3470	3320	3200	3110
60	3610	3430	3280	3150	3040	2960
62	3390	3240	3100	2990	2890	2810
64	3190	3060	2930	2820	2740	2670
220°F						
54	4410	4230	4050	3890	3750	3620
56	3990	3780	3600	3440	3300	3180
58	3700	3480	3280	3110	2970	2850
60	3430	3210	3030	2880	2760	2660
62	3150	2970	2800	2670	2570	2480
64	2900	2740	2590	2470	2380	2300

Although the five systems presented a satisfactory correlation within themselves the correlation should be applied with caution to other fields.

The basis for the 160°F data presented in Tbl. 11.2 is shown in Fig. 11.5.

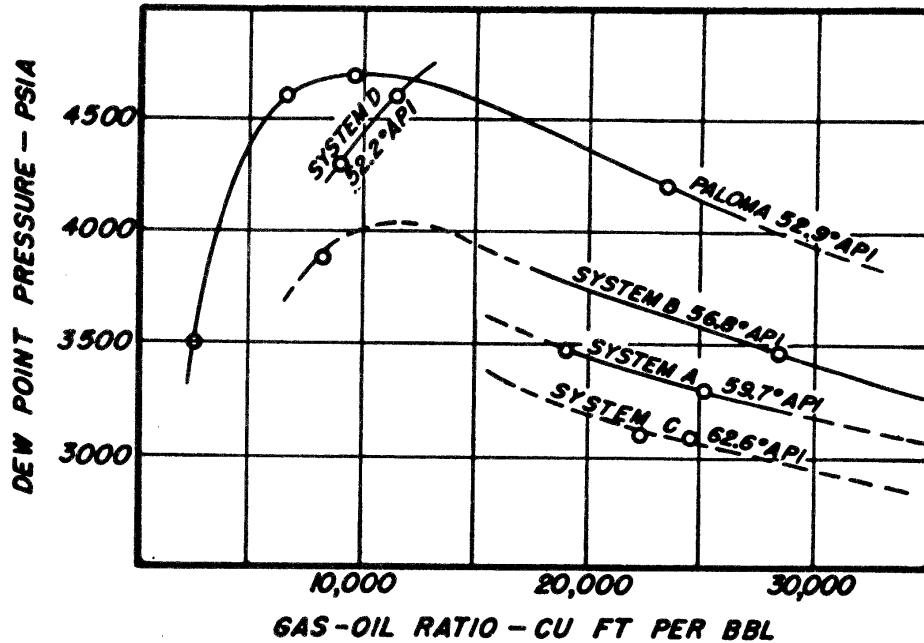


Fig. 11.5. Influence of Gas-Oil Ratio and Tank Oil Gravity upon Retrograde Dew Point Pressure at 160°F
(After Sage and Olds Trans. AIME 1947)

The curves illustrate a behavior of condensate systems the understanding of which is important in the correct application of laboratory results to field conditions. The fact that the curves show maximum dew point pressure values makes it possible for a laboratory system to have a dew point pressure equal to the reservoir pressure but not necessarily to have the same composition as the reservoir vapor. For example the uppermost curve shown in Fig. 11.5 exhibits a dew point pressure of 4700 psia over a gas-oil ratio range from 8000 to 12000 ft³/bbl. If the producing gas-oil ratio at the time of sampling were 8000 ft³/bbl and the reservoir pressure were 4700 psia the data could be interpreted as indicating the reservoir fluid to be at its dew point and to have the properties of the 8000 ft³/bbl system. On the other hand it is not inconceivable that the actual reservoir vapor might be the equivalent of the 12000 ft³/bbl system and that the 8000 ratio was a fortuitous result of simultaneous production of vapor and retrograde reservoir liquid into the well bore. Under such circumstances the use of the phase behavior of the 8000 ft³/bbl system to predict the reservoir fluid behavior could result in errors in both the correct formation volume of the reservoir system and the quantity of retrograde liquid that would be formed as a result of reservoir pressure decline.

Because of the domelike shape of the gas-oil ratio-dew point pressure curve one should not place as much emphasis on laboratory dew point pressure measurements as is usually placed on bubble point pressures. As will be shown later there is a monotonic relation of bubble point pressure with gas-oil ratio so that simultaneous flow of free gas and saturated liquid into the well bore will always cause the bubble point pressure of the recombined dissolved gas system to be higher than the bottom-hole pressure. On the other hand simultaneous flow of retrograde reservoir liquid and saturated vapor into the well bore may result in a condensate system having a dew point pressure lower than the bottom-hole pressure and thus cause one to believe the reservoir fluid to be undersaturated.

11.3.2 Liquid Phase Behavior.

The most characteristic curve of gas-condensate systems relates the quantity of equilibrium liquid phase to pressure. This curve usually is determined at reservoir temperature to indicate the hydrocarbon liquid saturation that will be formed in the reservoir sands as a result of pressure decline. These data are important to economic studies of cycling operations. Three type curves are shown in Fig. 11.6.

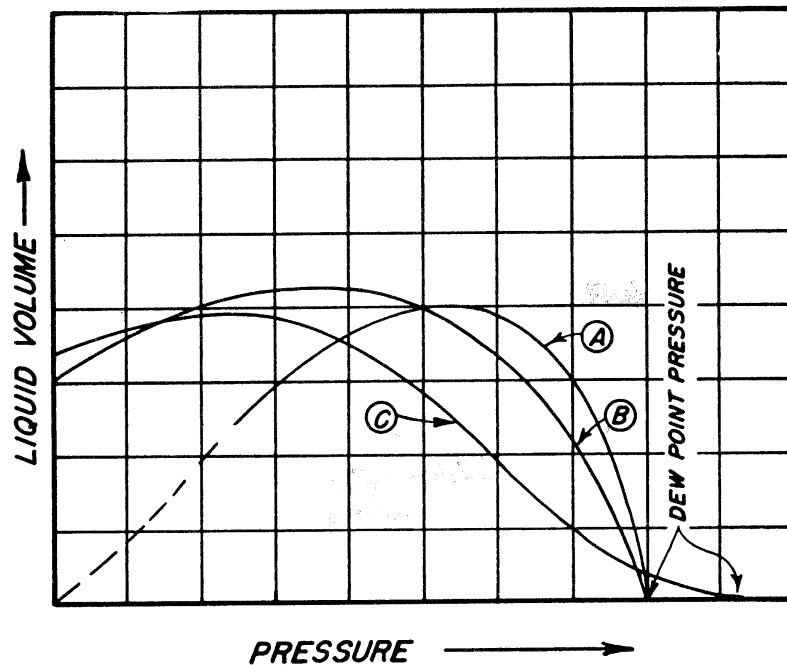


Fig. 11.6. Curves Illustrating Several Types of Liquid Behavior of Condensate Systems

The curve illustrated by A shows the relation between volume % liquid and pressure for a flash process; the approach to the origin is merely an expression of the large total volume necessary to reach a low pressure for a system of constant over-all composition.

The curves illustrated by B and C are characteristic of differential processes in which the total volume of the system is maintained constant during the pressure reduction. Temperature and the initial composition of the system will affect the dew point pressure and the quantity of liquid at any pressure. In general, a system that has a surface producing a gas-oil ratio of 15000 ft³/bbl will give a maximum liquid content of 4 to 6 % by volume at reservoir temperatures near 200°F whereas a 50000 ft³/bbl system ordinarily will give less than 1 % liquid. Curves B and C illustrate differences in liquid volume behavior near the dew point pressure. While insufficient data are available in the literature to explain these differences it is believed that the asymptotic approach to the dew point pressure of curve C is caused by a system that has a wider range of heavier components than a system that gives the B curve.

Where actual volumes of liquid formed during flash and differential processes have been compared it has been found that the two liquid quantities have been nearly equal during the first 25 % reduction in pressure. As further pressure reductions occur the differential process normally results in less liquid because portions of the heavier components are removed from the system as the differential process proceeds. In natural systems the maximum amount of liquid occurs in the pressure range 1200 to 2500 psi for both the flash and differential processes. At lower temperatures larger amounts of liquid are formed and maximum liquid occurs at a lower pressure.

11.3.3 Vapor Phase Behavior.

Just as the volume of liquid formed by a condensate system goes through a maximum near 2000 psi the quantity of heavy components held in the vapor phase goes through a minimum usually at about 1000 psia. This behavior of condensate systems is illustrated quite clearly by the results of the differential tests shown in Fig. 11.7.

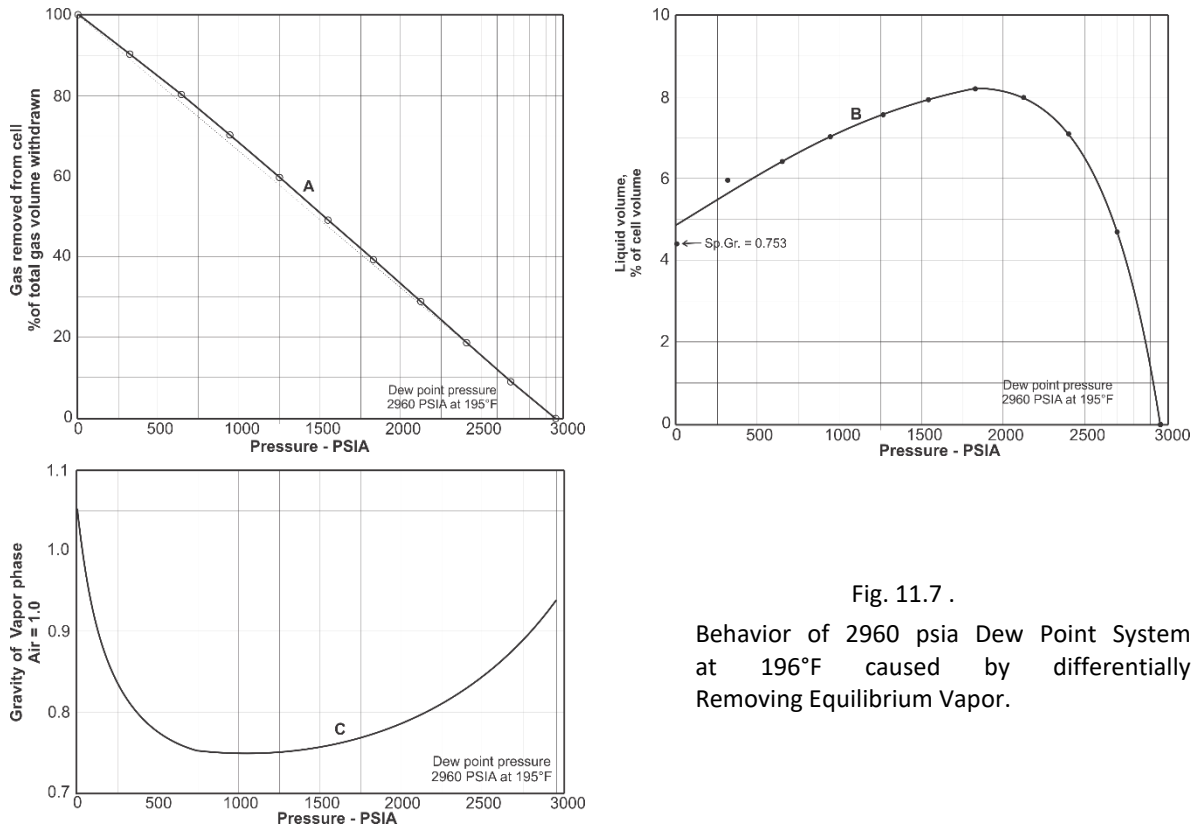


Fig. 11.7 .
Behavior of 2960 psia Dew Point System at 196°F caused by differentially Removing Equilibrium Vapor.

The percentage of the system that has been produced is related to the resulting decline in system pressure as shown by curve A of Fig. 11.7. For example, 20% of the system available in the vapor form between the dew point pressure of 2960 and 14.7 psia is produced during pressure decline from 2960 to 2370 psia. The relation can be used directly with field production data to calculate reserves where it is known that the reservoir HC space does not change because of reservoir pressure decline. Curves of the type illustrated are characteristically slightly S-shaped because of changes that occur in the compressibility factor and system composition as pressure declines.

Curve B of Fig. 11.7 shows the amount of retrograde liquid formed during the differential test.

An indication of the richness of the vapor phase at any pressure is given by curve C Fig. 11.7. The smooth curve represents the instantaneous value of the vapor phase gravity. As will be shown in the next section the calculated quantity of butanes plus material in the vapor phase decreases from 5.2gal/MCF at the dew point to 2.6gal/MCF at 1 000 psia.

11.3.4 Phase Compositions

Laboratory tests on gas-condensate systems usually are made to determine the volumetric behavior of the systems at reservoir and surface trapping conditions. For special engineering studies however, one may need to know the phase compositions at various pressures during the depletion of a condensate reservoir. The composition of each phase can be determined experimentally but to do so requires more work than often is believed justified.

In gas-condensate systems the calculated phase compositions and volumes in particular the liquid phase volume can be greatly in error through use of present known equilibrium constants. It becomes advisable therefore when making such calculations to control them by use of experimentally determined volume measurements.

The remaining portion of this chapter will discuss the steps required in such calculations. The example given pertains to an actual gas-condensate system that has a dew point pressure of 2 960 psia at 195°F and the composition shown in Example 11.4. The results of laboratory differential tests to determine the volume of liquid phase and the pressure reduction-vapor withdrawal relation have been given previously in Fig. 11.7.

As a starting point in the calculation let it be assumed that a reservoir hydrocarbon space of 100 ft³ represents the unit working volume of the system in which we are interested. It will be assumed that this volume remains constant throughout the calculation process. The calculations in Example 11.4 determine that 54.0 moles constitute the original system at 2960 psia and 195°F.

Example 11.4.

Calculation of the Original and Final Quantities of Material Remaining in the Unit Reservoir Volume

Basis (100 ft³ hydrocarbon space)

Initial dew point pressure	2960 psia
Final pressure	14.7 psia
Temperature	195°F

Initial condition at 2960 psia:

1	2	3	4	5	6
Comp.	Mole Frac. z	Critical constant of comp.		Pseudo critical T and P	
		T _c	P _c	z·T _c	z·P _c
C1	0.7527	344	673	258.7	506.0
C2	0.0766	550	709	42.1	54.3
C3	0.0441	666	618	29.4	27.2
C4	0.0309	750	540	23.2	16.7
C5	0.0221	838	484	18.5	10.7
C6	0.0206	915	434	18.8	8.9
C7+	0.0530*	1042	418	55.3	22.2
Σ	1.0000			446.0	646.0

*C7+ molecular weight = 114

C7+ specific gravity = 0.755

$$pP_r = \frac{2960}{646.0} = 4.58$$

$$pT_r = \frac{655}{446.0} = 1.47$$

giving

$$Z = 0.78 \text{ (Fig. 8.4)}$$

$$n = \frac{PV}{ZRT} = \frac{2960 \cdot 100}{0.78 \cdot 10.73 \cdot 655} = 54.0 \text{ lb moles}$$

Final condition at 14.7 psia:

Specific gravity of cell residue:

$$\gamma = 0.753 \text{ (60/60 }^\circ\text{F) (Fig. 11.7)}$$

From Fig. 9.4 (Liquid Chapter);

density of cell residue (14.7 psia and 195°F)

$$= (0.753 \cdot 62.37) - 3.6 = 43.4 \text{ lb/ft}^3$$

From Fig. 11.3; assumed molecular weight (M) residue = 113.

From Fig. 11.7 volume of residue liquid (14.7 psia and 195°F) = 4.9 ft³

$$\text{Quantity of residue liquid;} = \frac{4.9 \cdot 43.4}{113} = 1.88 \text{ lb moles}$$

$$\text{Quantity of residue gas} = \frac{(100-4.9)}{379} \cdot \frac{520}{655} = 0.20 \text{ lb moles.}$$

Total quantity of residue gas and liquid = 1.88 + 0.20 = 2.08 lb moles

Material removed from the unit volume going from 2960 to 14.7 psia = 54.0 - 2.1 = 51.9 lb moles

The calculations also show that 2.1 lb moles of material remain in the 100 ft³ volume at 14.7 psia and 195°F. The difference between this value and the original (51.9 lb moles) represents the quantity of material removed during the differential process.

The second step is to determine a set of equilibrium constants with which to calculate phase compositions. An immediate test of the applicability of any set of constants is that the sum of $\frac{n_k}{K_k}$ must be equal to unity (Eq. 10.18 – Phase equilibrium chapter) at the observed dew point pressure of 2 960 psia. In Example 11.5 are presented three dew point pressure calculations using Roland Smith and Kaveler's constants for 200°F. (It is not justified at this point to work out constants corrected to 195°F.)

Example 11.5.

Determination of Relative Equilibrium Constants to Compute Dew Point Conditions.

Solution :

Basis: Roland Smith and Kaveler's Data 200°F

$$\sum_{k=1}^{k=j} \frac{n_k}{K_k} = 1$$

1	2	3	4	5	6	7	8
Comp.	Mole Frac. n	Pressure 2960 psia		Pressure 3500 psia		Pressure 3600 psia	
		K	$\frac{n}{K}$	K	$\frac{n}{K}$	K	$\frac{n}{K}$
C1	0.7527	1.67	0.4500	1.35	0.5575	1.30	0.5795
C2	0.0766	0.92	0.0830	0.96	0.0799	0.97	0.0790
C3	0.0441	0.69	0.0649	0.82	0.0538	0.85	0.0519
C4	0.0309	0.52	0.0594	0.71	0.0435	0.76	0.0407
C5	0.0221	0.34	0.0650	0.54	0.0410	0.60	0.0379
C6	0.0206	0.23	0.0896	0.44	0.0468	0.50	0.0412
C7+	0.0530	0.095	0.5580	0.28	0.1892	0.35	0.1514
	1.000		1.3690		1.0117		0.9816

Dew Point Pressure P_d = 3540 psia.

These constants satisfy the condition that the summation of $\frac{n_k}{K_k}$ is equal to unity only at 3540 psia. As the system under study gave an observed dew point at 2960 psia Roland's 3540 psia values must be used at 2960 psia pressure. Thus the published constants are used here primarily for their relative values and not for their absolute values.

The methods presented in the Vapor - liquid equilibrium chapter are used next to estimate equilibrium constants at pressures below the dew point pressure. Assuming an apparent convergence pressure of the system several hundred psi above the dew point pressure makes it possible to calculate by use of Fig. 10.4 and Fig. 10.5 (K-values) the minimum K's and the pressure at which they occur. The smooth curves shown in Fig. 11.8 have been drawn to connect the pressures at which the individual constants are equal to unity (vapor pressure) the minimum constants the constants at the experimental dew point pressure and a convergence pressure of 3500 psia. Roland Smith and Kaveler's curves at 200°F also have been plotted in Fig. 11.8 to aid in drafting the 195°F curves. The curve constructed through the solid points will be explained later.

The fourth step is to establish a calculation procedure to simulate the differential process of producing equilibrium vapor from the unit volume. A simple stepwise calculation process is diagrammed in Fig. 11.9.

This calculation process can be visualized best as taking place in a closed vessel in which it is possible to remove and inject an inert liquid such as mercury into the bottom of the vessel. Equilibrium vapor phase material can be removed from the top of the vessel. Referring to Fig. 11.9 condition I pictures 54.0 lb moles of original dew point material at 2960 psia. Condition II pictures the removal of 5.19 moles of vapor from the vessel at a constant pressure of 2960 psia leaving a total of 48.81 lb moles of material in the system at condition III. In going from condition II to condition III the exact amount of mercury previously injected into the vessel is withdrawn so that the total volume of the system at condition III again is 100 ft³. The volume that the liquid phase occupies at 2650 psia is read from curve B Fig. 11.7. At 2650 psia a second unit of 5.19 moles of equilibrium vapor is removed from the vessel. The second modified system pictured by IV then is brought to the original 100 ft³ volume at 2370 psia as pictured by condition V. The process then is repeated at pressures of 2090 1800 1520 1230 940 640 320 and 15 psia. These particular pressures are read from curve A Fig. 11.7 and represent the pressures which result from withdrawal of 10 % increments or 5.19 lb moles of equilibrium vapor phase.

An alternate process can be pictured in which the 5.19 lb moles are removed from the vessel after pressure reduction rather than before. The composition of the vapor and liquid phases calculated by this second process does not differ materially from that calculated by the process just described. Of course, the greater the number of steps the closer the two methods will agree with each other and for an infinite number of steps each process becomes a true differential process.

The PVT data presented in Fig. 11.7 are in terms of volumes whereas equilibrium constant calculations make use of molal phase quantities. To convert the cubic feet of liquid phase calculated from Fig. 11.7 to pound moles of liquid it is necessary to know the molecular weight and density of the liquids as they exist at the elevated pressures and 195°F. These values can be determined by the methods presented in the Liquid chapter provided the pressure-density and pressure-molecular weight relations of the heptanes plus fraction are known. The only direct laboratory data available are the molecular weight and specific gravity of the total heptanes plus in the original system and the specific gravity of the residual liquid remaining in the laboratory cell at the completion of the differential process. These data have been used with the trend of the curves in Fig. 10.2 to prepare the assumed pressure relations shown in Fig. 11.10.

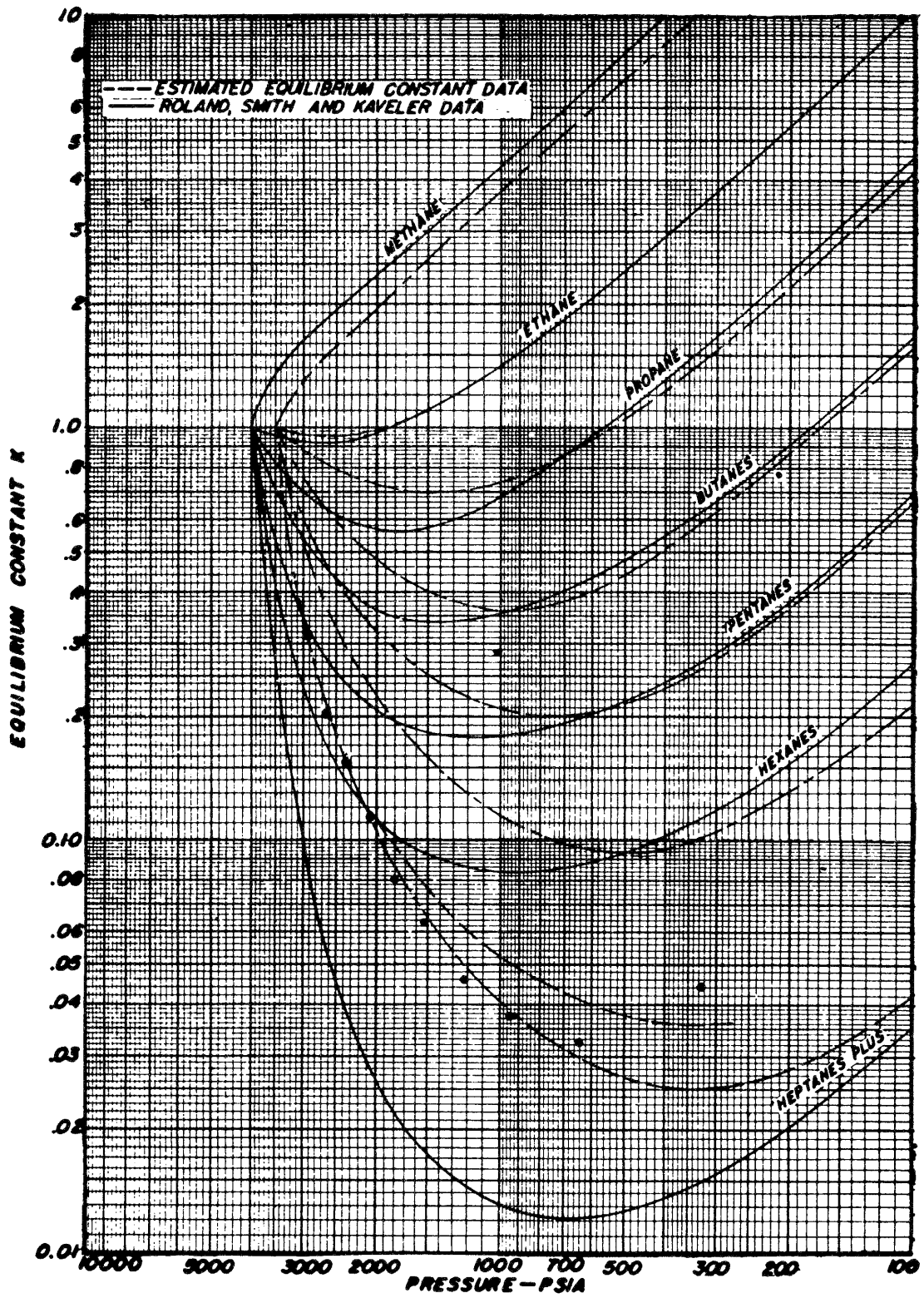


Fig. 11.8 . Estimated Equilibrium Constant Data from Roland Smith and Kaveler Data at 200°F

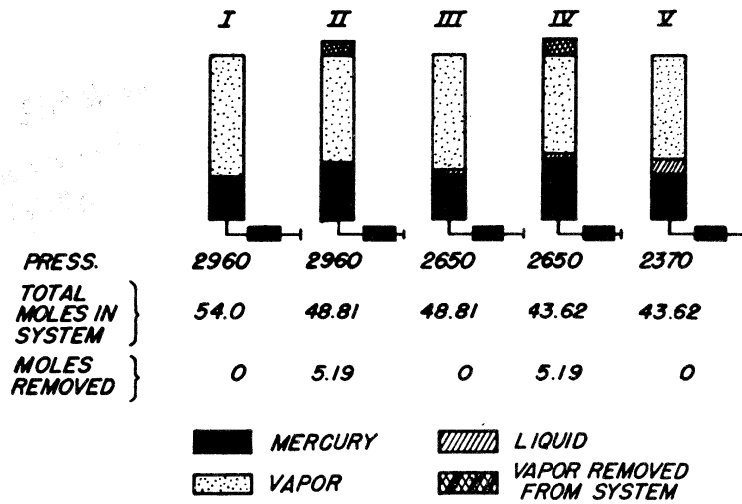


Fig. 11.9 . Diagram Illustrating Stepwise Calculation Process

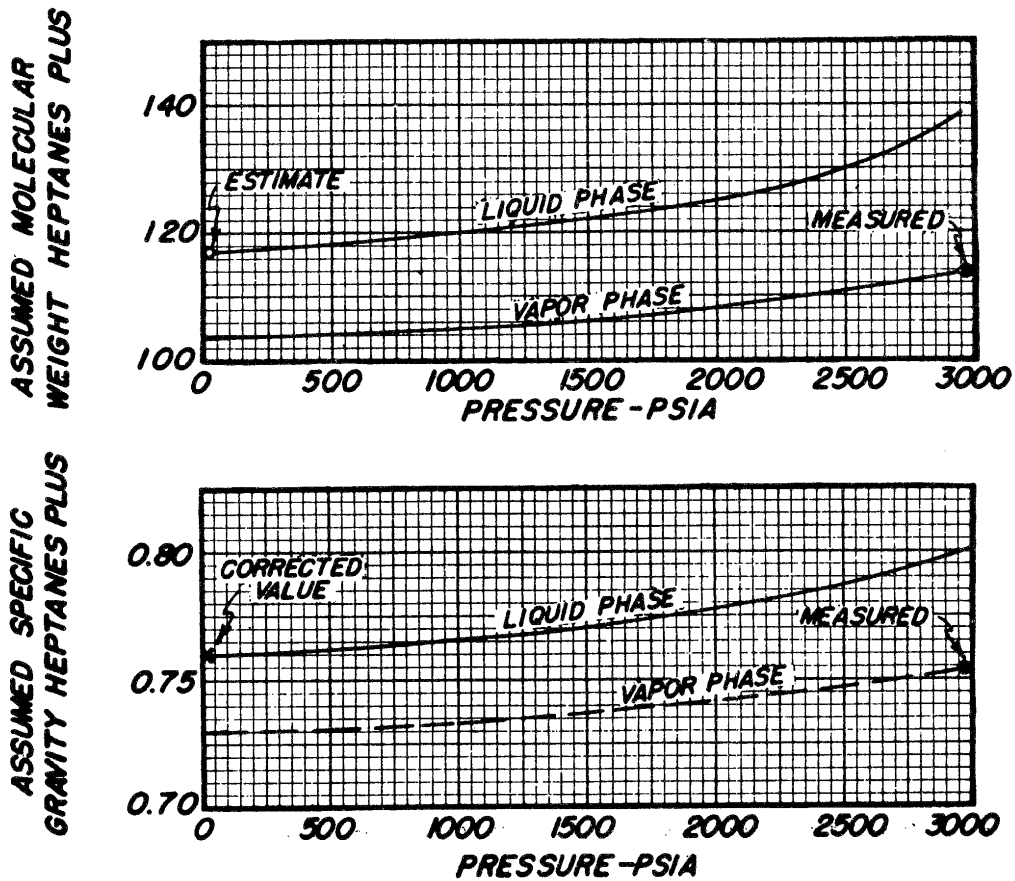


Fig. 11.10 . Assumed Relationship of Molecular Weight and Specific Gravity of Heptanes Plus to Pressure

The measured values of 114 and 0.755 are shown as corresponding to the dew point vapor. Inasmuch as the residual cell liquid at 14.7 psia and 195°F is of the order of 85 % heptanes plus material its measured specific gravity of 0.753 has been corrected upward to correspond to a heptanes plus specific gravity of 0.760. From this gravity the molecular weight of the heptanes plus in the residue liquid is estimated to be 117. Unfortunately, these are all the data available from which to draw the

curves shown in Fig. 11.10. Whether the curves are straight lines or have curvature as shown can only be surmised. The heptanes plus liquid density at 2960 psia is an assumed value because the density cannot be calculated from the existing data. In short, Fig. 11.10 illustrates one of the great needs in vaporization calculation work: the behavior of the heptanes plus fraction at different pressures.

The composition of the dew point liquid at 2960 psia and 195°F is shown in Example 11.6.

The method of obtaining the equilibrium constants listed in column 3 has been discussed previously in connection with Fig. 11.8. The properties of the heptanes plus fraction required to calculate its volume and weight were obtained from Fig. 11.10. Volumes of the individual components listed in column 6 were obtained by use of the atmospheric liquid densities (tabled values). An arithmetic average of the liquid densities of iso and normal butane and pentane was used in the calculation. It is interesting to note that almost 20 weight% of the dew point liquid is methane.

Example 11.6.

Calculation of Properties of Dew Point Liquid at $(P\&T)_{res}$, 2960 Psia and 195 °F,

Solution:

Basis: One pound mole of liquid.

1	2	3	4	5	6
Comp	Mole Frac. n	K at $(P\&T)_{res}$	Liq.Phase Comp. $\frac{n}{K}$	Weight lb/ lb mole liquid $\frac{n \cdot M}{K}$	Liq.Vol. at SC, ft ³ $\frac{n \cdot M}{K \cdot \rho}$
C1	0.7527	1.32	0.5720	9.15	
C2	0.0766	0.96	0.0799	2.40	
C3	0.0441	0.83	0.0531	2.34	0.074
C4	0.0309	0.74	0.0418	2.43	0.068
C5	0.0221	0.57	0.0388	2.80	0.071
C6	0.0206	0.47	0.0438	3.77	0.091
C7+	0.0530	0.31	0.1706	23.55*	0.472*
	1.0000		1.0000	46.44	0.776

SC = 14.7 psia & 60°F

$(P\&T)_{res}$ = 2960 Psia and 195 °F

*Assumed molecular weight of C7+ = 138 (Fig. 11.10)

Assumed specific gravity of C7+ (Fig. 11.10):

$$\gamma_{C7+} = 0.800$$

$$\text{Density propane plus at 14.7 psia and 60°F} = \frac{35.45}{0.776} = 45.7 \text{ lb/ft}^3$$

$$\text{Weight\% ethane in ethane plus} = \frac{2.40}{37.63} = 6.3 \%$$

$$\text{Weight\% ethane in ethane plus} = \frac{9.15}{46.44} = 19.5 \%$$

From Fig. 9.2 pseudo-liquid density at 14.7 psia and 60°F = 31.3 lb/ft³

From Fig. 9.3 density at 2960 psia and 60°F = 31.3 + 2.7 = 34.0 lb/ft³

From Fig. 9.4 density at $(P\&T)_{res}$ = 34.0 - 9.3 = 24.7 lb/ft³

$$\text{Pound moles liquid per cubic foot liquid} = \frac{24.7}{46.44} = 0.532$$

The calculation of vapor and liquid compositions at 2650 psia is shown in Example 11.7. In part B the number of moles of liquid and vapor in equilibrium is calculated from the total moles (48.81)

comprising the first modified system and the quantity of liquid phase determined by the laboratory tests. To convert the measured liquid volume to pound moles it is assumed that the factor pound moles per cubic foot evaluated at 2960 psia (Example 11.6) also will apply at 2650 psia. This assumption is checked and if necessary is corrected after the composition of 2650 psia liquid has been calculated in part B.

Part B in Example 11.7 outlines the vaporization calculation at 2650 psia using the relative values of L and V determined in part A. The following procedure has been followed in making these calculations:

- (1) With the exception of heptanes plus the mole fraction of each component in the liquid phase is calculated by the usual methods.
- (2) The mole fraction of C7+ in the liquid phase (0.2059) is set at the value necessary for $\sum x = 1$.
- (3) Working backward from the values of x_{C7+} and n_{C7+} the values of $L + V \cdot K_7$ are calculated.
- (4) The mole fraction of heptanes plus in the vapor phase y_{7+} is calculated from the values of K_{C7+} and x_{C7+} using the relation $y = K \cdot x$.
- (5) The mole fractions of the components in the vapor phase are made to add to unity by minor adjustments in the mole fraction of the methane.

Example 11.7.

Calculation of Vapor and Liquid Composition at 2650 Psia

Part A:

Estimation of Vapor to Liquid Ratio (Basis: 48.81 lb moles of system)

From calculations in Example 11.6, assume that the liquid phase will have the equivalent of 0.53 lb moles per cu ft.

From Fig. 11.7; volume of liquid at 2650 psia is 5.2 cu ft.

Therefore, moles of liquid = 5.2 · 0.53 = 2.76.

Mole fraction liquid, $L = \frac{2.76}{48.8} = 0.056$; Mole fraction vapor, $V = 0.944$

Part B:

Calculation of Vapor and Liquid Composition (Basis: One pound mole of system)

1	2	3	4	5	6	7
Comp.	System Mole Fraction	Eq.const K at P and T 2650psia, 195°F	L = 0.056 V = 0.944		Liq.Phase Mole frac. x	Vap.Phase Mole frac. y
			V·K	L+V·K		
C1	0.7527	1.48	1.397	1.453	0.5180	(0.7659)
C2	0.0766	0.95	0.897	0.953	0.0804	0.0764
C3	0.0441	0.82	0.774	0.830	0.0531	0.0435
C4	0.0309	0.63	0.595	0.651	0.0475	0.0299
C5	0.0221	0.47	0.444	0.500	0.0442	0.0208
Ce	0.0206	0.37	0.349	0.405	0.0509	0.0188
C7+	0.0530	(0.217)	(0.205)	(0.261)	(0.2059)	0.0447
Σ	1.0000				1.0000	1.0000

Part C:

Properties of Liquid Phase (Basis: One pound mole of liquid)

1	2	3	4
Comp.	Liq.Phase Mole frac. x	Weight of Comp. $x \cdot M$	Liq.Volume at P and T 14.7psia, 60°F $\frac{x \cdot M}{\rho}$
C1	0.5180	8.31	
C2	0.0804	2.42	
C3	0.0531	2.34	0.074
C4	0.0475	2.76	0.076
C5	0.0442	3.19	0.082
Ce	0.0509	4.39	0.106
C7+	0.2059	27.18*	0.552
Σ	1.0000	50.59	0.890

*From Fig. 11.10; Molecular weight C7+ = 132 , Specific gravity C7+ = 0.790

Density propane plus at 14.7 psia and 60°F (Fig.9.2)

$$\rho_{C3+} = \frac{39.86}{0.890} = 44.8 \text{ lb/ft}^3$$

Weight per cent ethane in ethane plus

$$Wt\% C2 = \frac{2.42}{42.29} = 5.7$$

Weight per cent methane in system

$$Wt\% C1 = \frac{8.31}{50.59} = 16.6$$

From Fig.9.2, pseudo-liquid density at 14.7 psia and 60°F = 33.2 lb/ft³

From Fig. 9.3, density at 2650 psia and 60°F = 33.2 + 2.0 = 35.2 lb/ft³

From Fig. 9.4, density at 2650 psia and 195°F = 35.2 - 7.9 = 27.3 lb/ft³

Lb moles liq./ ft³ liq. = $\frac{27.3}{50.59} = 0.539$ (Assumed value of 0.53 in Part A is satisfactory.)

Part D:

Calculation of modified system at 2370 psia (Basis: 48.81 lb moles of material at 2650 psia).

1	2	3	4	5	6	7
Comp	First modified System				Second modified System	
	2650 psia System		2650 psia System		2370 psia System	
	Mole frac.	lb moles	Mole frac.	lb moles	Mole frac.	lb moles
C1	0.7527	30.740	0.7659	3.975	32.765	0.7511
C2	0.0766	3.739	0.0764	0.396	3.343	0.0766
C3	0.0441	2.152	0.0435	0.226	1.926	0.0442
C4	0.0309	1.508	0.0299	0.155	1.353	0.0310
C5	0.0221	1.079	0.0208	0.108	0.971	0.0223
C6	0.0206	1.005	0.0188	0.098	0.907	0.0208
C7+	0.0530	2.587	0.0447	0.232	2.355	0.0540
Σ	1.0000	48.810	1.0000	5.190	43.620	1.0000

5.19 lb moles of 2650 psia vapor removed.

The procedure just outlined assumes that all equilibrium constants are correct except that of the heptanes plus fraction. This assumption of course may not be strictly valid but it presents a workable

method of calculating the phase compositions. The net effect of throwing all equilibrium constant errors into the heptanes plus constant is shown by the solid points in Fig. 11.8. If the methane constant curve had been drawn slightly higher so that its value at 2650 psia was 1.60 rather than 1.48 the calculated equilibrium constant of the heptanes plus fraction would have been 0.173 rather than 0.217. This would have caused better agreement between the calculated heptanes plus curve and the curve previously obtained by consideration of the apparent convergence pressure. However unless a much wider discrepancy is obtained than that illustrated in Fig. 11.8 it usually does not pay to undertake further refinement of the curve.

The calculations in part C of Example 11.7 are required to check the assumed value of 0.53 lb moles/ft³ of liquid used in part A. The agreement of 0.53 with 0.536 is considered satisfactory. Had a check not been obtained, it would have been necessary to recalculate parts A, B, and C.

The calculations in part D of Example 11.7 correspond to the removal of 5.19 lb moles of 2650 psia equilibrium vapor, as pictured by condition IV of Fig. 11.9. The remaining moles of material constitute the second modified system that will be expanded to 2370 psia.

The results of the ten-step calculation process are shown graphically in Fig. 11.11 and Fig. 11.12.

The composition, density, and molecular weight of the liquid phase at various pressures are shown in Fig. 11.11. As might be expected, the quantity of dissolved methane continuously decreases as the pressure on the system is reduced. Likewise, the proportion of heavy components increases with decreasing pressure. Intermediate components tend to have maximum concentration in the liquid phase at pressures near the pressure of their minimum equilibrium constant. For example, the maximum butane concentration in the liquid is near 850 psia, and the equilibrium constant of butanes (Fig. 11.8) is a minimum at that pressure. The behavior of the vapor phase is shown in Fig. 11.12. It will be noted that the calculated minimum gravity of 0.780 at 1000 psia does not check the laboratory test value of 0.75 shown in Fig. 11.7. This may be caused in part by the high value of the heptanes plus equilibrium constant. A second factor that may contribute to the discrepancy is the assumed molecular weight of the heptanes plus vapor.

It can be appreciated from the foregoing section that calculated compositions of the vapor and liquid phases in a gas-condensate system may be in error because of any one of the numerous assumptions that have to be made in the calculation process. The method outlined above, where the volumetric behavior of the system determined experimentally in the laboratory is used as a control in the calculation process, is believed to be the most satisfactory one available. The accuracy of these calculations can be expected to improve as additional data become available on the behavior of condensate systems.

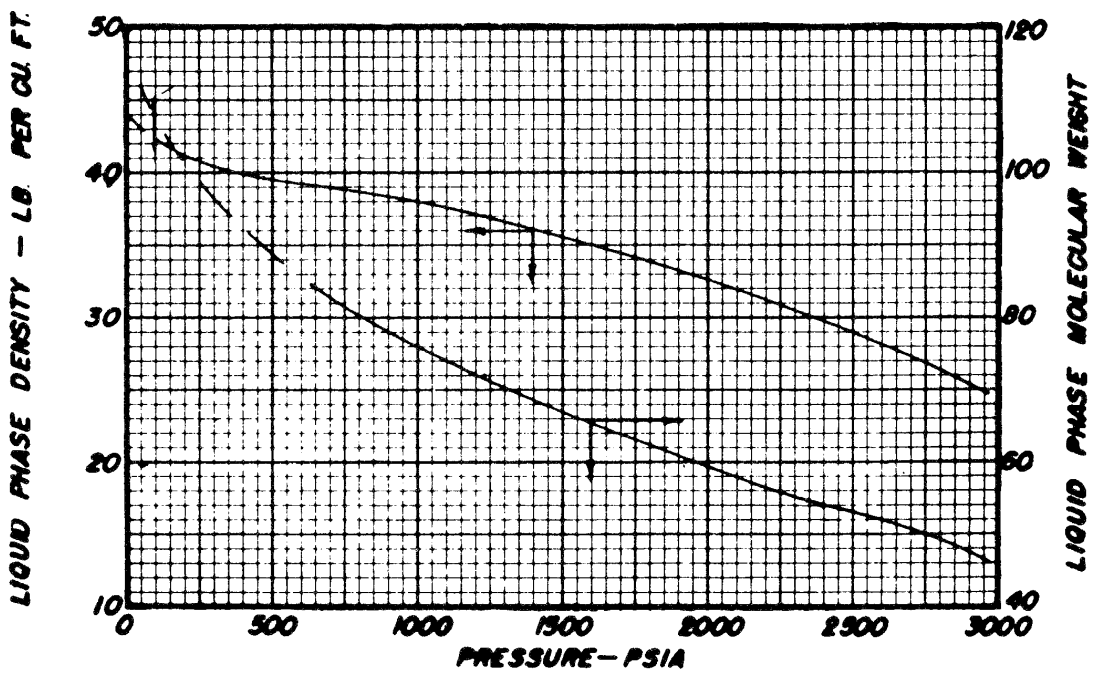
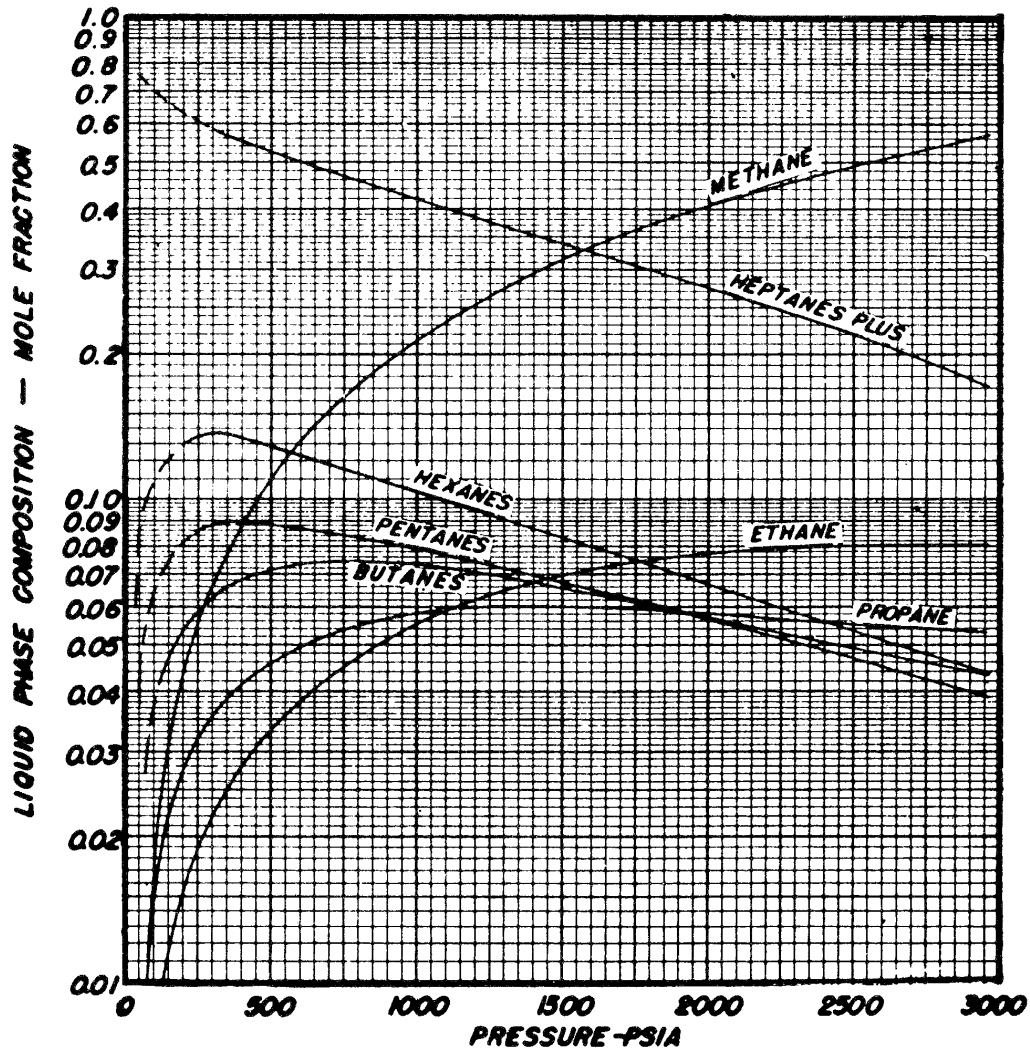


Fig. 11.11 . Calculated Composition and Properties of Liquid Phase

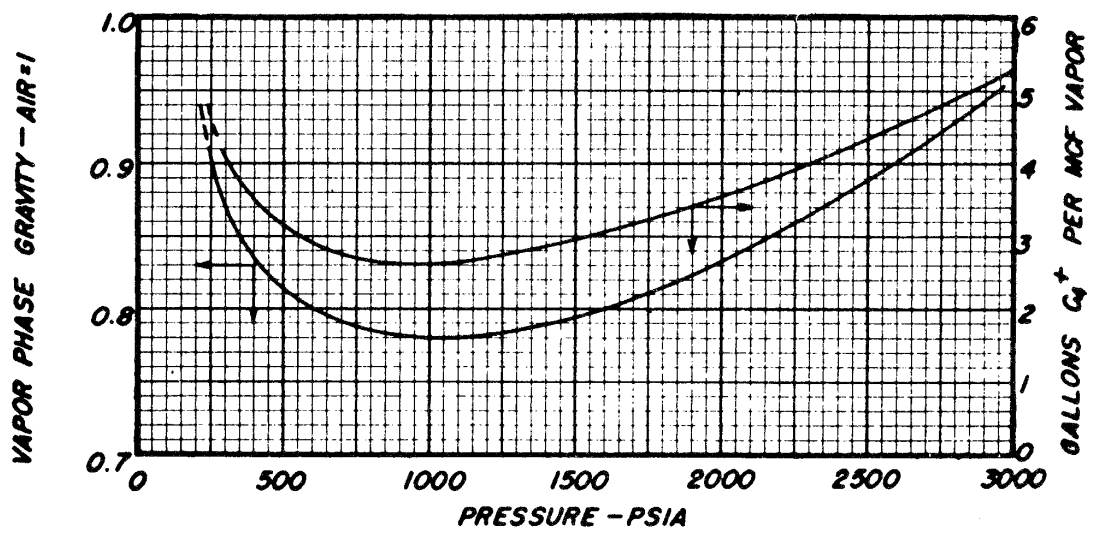
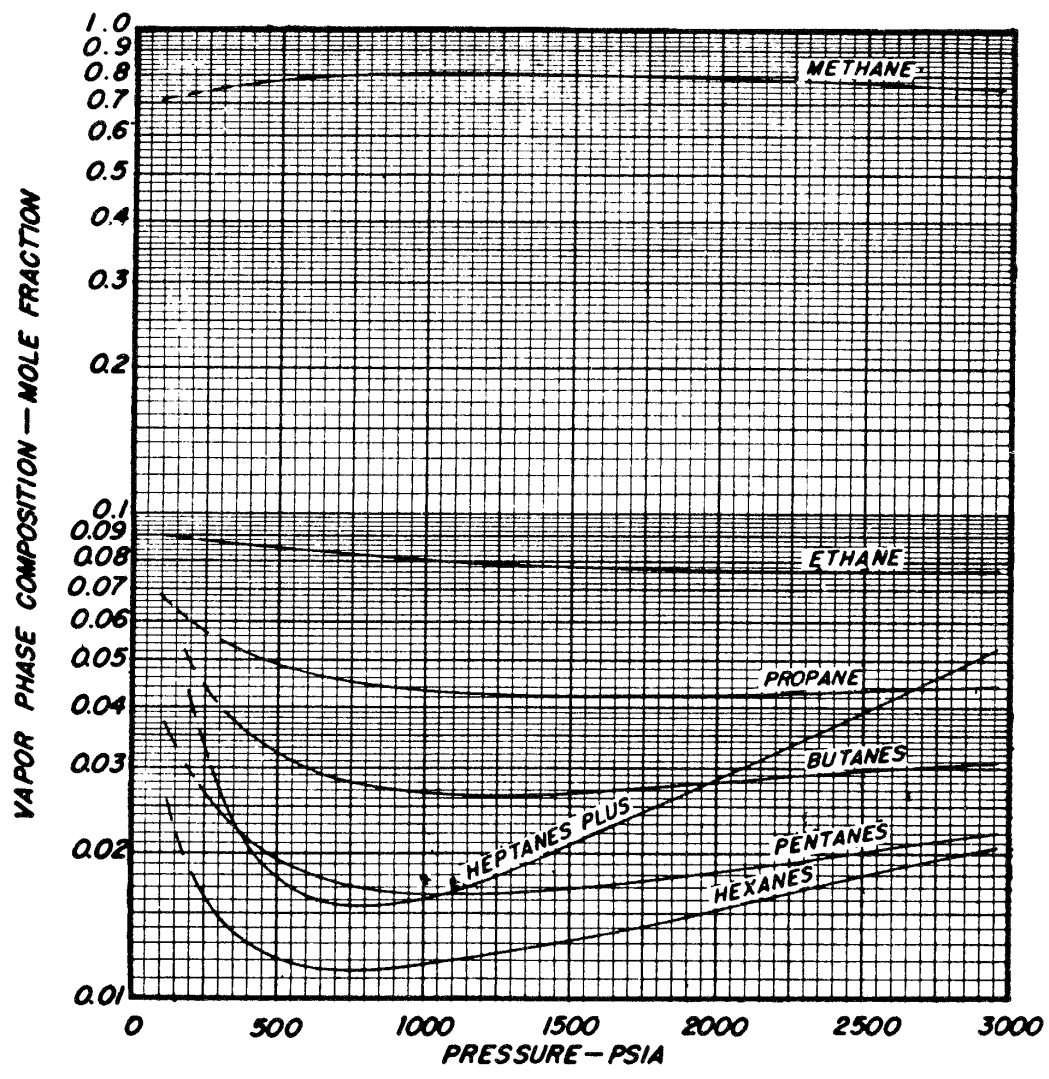


Fig. 11.12. Calculated Composition and Properties of Vapor Phase

11.4 Recovery calculations based on CVD analysis at $P_{res} < P_d$.

At reservoir pressures above P_d , recovery calculations for gas condensates could be performed as described in the Gas chapter.

Recovery calculations below P_d are more complicated due to the presence of both a gas phase and a liquid phase in the reservoir. The production could be estimated based on the Constant Volume Depletion experiments performed in the laboratory on reservoir fluid, where we have an immobile liquid phase. The calculation is done by scaling up the recovery from the PVT-cell at each pressure step. At constant temperature (normally = T_{res}) and with a closed reservoir (i. e. HCPV = constant), the recovery from the reservoir at pressure step j at P_j could be expressed as:

$$\frac{(\Delta G_p)_j}{(HCPV)} = \frac{(\Delta V_g)_{j SC}}{(V_{cell})_d} \quad (11.7)$$

where:

$(\Delta G_p)_{j SC}$: Gas volume of wellstream at standard conditions (SC)

$(HCPV)$: Hydrocarbon pore volume

$(\Delta V_g)_{j SC}$: Standard volume of gas taken out of the PVT cell at pressure P_j

$(V_{cell})_d$: Volume of fluid in the PVT cell at P_d

$$(\Delta G_p)_j = (HCPV) \frac{(\Delta V_g)_{j SC}}{(V_{cell})_d} = \frac{P_d(HCPV)}{(Z_g)_d RT_{res}} \cdot V_m \cdot \frac{\frac{P_j(\Delta V_g)_{j SC}}{(Z_g)_j RT_{res}} V_m}{\frac{P_d(V_{cell})_d}{(Z_g)_d RT_{res}} V_m} \quad (11.8)$$

and

$(Z_g)_j$: Z-value of wellstream gas produced in pressure step j at P_j

$(Z_g)_d$: Z-value of reservoir gas at P_d

V_m : molar volume of gas at SC

T_{res} : Reservoir temperature (absolute temperature.)

After Shortening

$$(\Delta G_p)_j = \frac{(HCPV) P_j}{RT_{res} (Z_g)_j} \cdot \frac{(\Delta V_g)_j}{(V_{cell})_d} \cdot V_m \quad (11.9)$$

The number of wellstream moles produced in pressure step j :

$$(\Delta n_p)_j = \frac{(\Delta G_p)_j}{V_m} = \frac{(HCPV) P_j}{RT_{res} (Z_g)_j} \cdot \frac{(\Delta V_g)_j}{(V_{cell})_d} \quad (11.10)$$

Volume of Gas (SC) and STO from separators at pressure step j .

Calculation of volume gas from separators at SC in pressure step j , $(\Delta V_g)_{j sep}$, and volume STO, $(V_{STO})_j$ could be performed if we suppose the wellstream is separated as gas at SC with composition C1, C2, C3, i-C4, and n-C4 and STO has the composition of C5+, i. e. i-C5, n-C5, C6, C7,,C10+.

The composition of wellstream at pressure step j is $(y_i)_j$.

$$(V_g)_{j \text{ sep}} = (\Delta n_g)_j \cdot (y_{C1} + y_{C2} + \dots + y_{n-C4}) \cdot V_m \quad (11.11)$$

In order to calculate the volume of *STO*, we have to use the additive volume method, i. e. the liquid volume $V = \frac{m}{\rho} = \frac{n \cdot M}{\rho}$ of each of the components must be determined by :

$$(V_{STO})_j = V_{i-C5} + V_{n-C5} + V_{C6} + \dots + V_{C10+} \quad (11.12)$$

$$(V_{STO})_j = \frac{(\Delta n_g)_j (y_{i-C5}) M_{i-C5}}{\rho_{i-C5}} + \dots + \frac{(\Delta n_g)_j (y_{C10+}) M_{C10+}}{\rho_{C10+}}$$

where:

M_i : molecular weight of component i

ρ_i : liquid density of component i

The calculated data are put into a table where also the cumulative recovery of wellstream, $(V_g)_{well \ SC}$, separator gas, $(V_g)_{sep \ SC}$, and volume *STO*, (V_{STO}) :

j	P_j	$(\Delta n_g)_j$	$(\Delta G_p)_j$	$(V_g)_{j \text{ sep}}$	$(V_{STO})_{j \text{ sep}}$	$(V_g)_{well \ SC}$ cum	$(V_g)_{sep \ SC}$ cum	V_{STO} cum
0	P_d							
1	P_1							
2	P_2							
3	P_3							
Etc.	Etc.							

where the cumulative production is:

$$(V_g)_{well \ SC} = \sum (\Delta G_p)_j$$

$$(V_g)_{sep \ SC} = \sum (V_g)_{j \text{ sep}} \quad (11.13)$$

$$V_{STO} = \sum (V_{STO})_{j \text{ sep}}$$

PS!! Recovery calculation based on the above method is only valid when S_{or} obtained from core data is greater than the maximum liquid drop out during the CVD analysis.

11.5 Summary of Principal Points

Gas-condensate systems are identified by their high gas-oil ratios and lightly colored, high-gravity tank oils. The critical temperature of the naturally occurring condensate systems is lower than the temperature of the reservoir in which they are found. The systems normally exhibit retrograde behavior at pressures above 2000 psia because of the relation of the critical temperature of the system to the reservoir temperature.

The volumetric behavior of gas-condensate systems can be calculated easily by using the pseudocritical constant method proposed by W. B. Kay.[8] In many instances the pseudo-critical constants of a series of gas-condensate systems can be plotted against the gravity of the system; calculation charts then can be constructed that will give the volume of the system as a function of pressure, temperature, and system gravity. Fig. 11.3 presents an easy way of calculating the gravity of a well fluid from the quantities and gravities of the resulting surface gases and tank oil.

Naturally occurring gas-condensate systems, in general, contain greater than 75 mole per cent methane. The volume of the liquid phase in equilibrium with vapor under reservoir conditions seldom is greater than 10 per cent of the total volume occupied by the system. At pressures of the order of 3000 psia or greater, the retrograde liquid in a reservoir may contain up to 40 per cent liquid methane by volume. The density of the liquid under these conditions often is as low as 25 lb/ft³.

The quantity of butane plus material held in the vapor phase usually is at a minimum near 1000 psia. This means that production from high pressure condensate fields should follow an increasing gas-oil ratio trend until the reservoir pressure declines to near 1000 psia. The producing gas-oil ratio normally will decrease at pressures less than 1000 psia.

Calculation of phase compositions is made best by using experimentally determined phase volumes to control the equilibrium calculations. However, it should be recognized that the calculated compositions can be in error because of using non-representative equilibrium constants and incorrect assumptions regarding the pressure-density and pressure-molecular weight relations of the heptanes plus fractions.

11.6 References

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