

FIG. 4-80. Comparison of equilibrium ratios at 100°F for 1,000- and 5,000-psia convergence pressure. (From NGAA.¹³)

Hachmuth⁴¹ and Roland, Smith, and Kaveler⁴² are plotted in Fig. 4-81. The data of Katz are preferred for crude-oil systems, and the data of Roland et al. are preferred for condensate fluids.

Calculating Procedures for Solution of Equilibrium Problems. The basic relations developed in Eqs. (4-74) through (4-77) apply to hydrocarbon systems which exist at pressures and temperatures within the two-phase region. Equations (4-78) and (4-79) apply to the special conditions existing at the bubble-point and dew-point pressures.

The solution of the equations to determine the dew-point and bubble-point pressures requires the selection of appropriate equilibrium ratios at the temperature of interest. An estimate of the pressure is made, and the

K values so determined used in Eq. (4-81) or (4-82) depending on whether a bubble point or dew point is sought. If the sum of the calculated values does not equal 1, another pressure is selected and the calculations repeated. Trials are continued until the sum is determined within the desired accuracy.

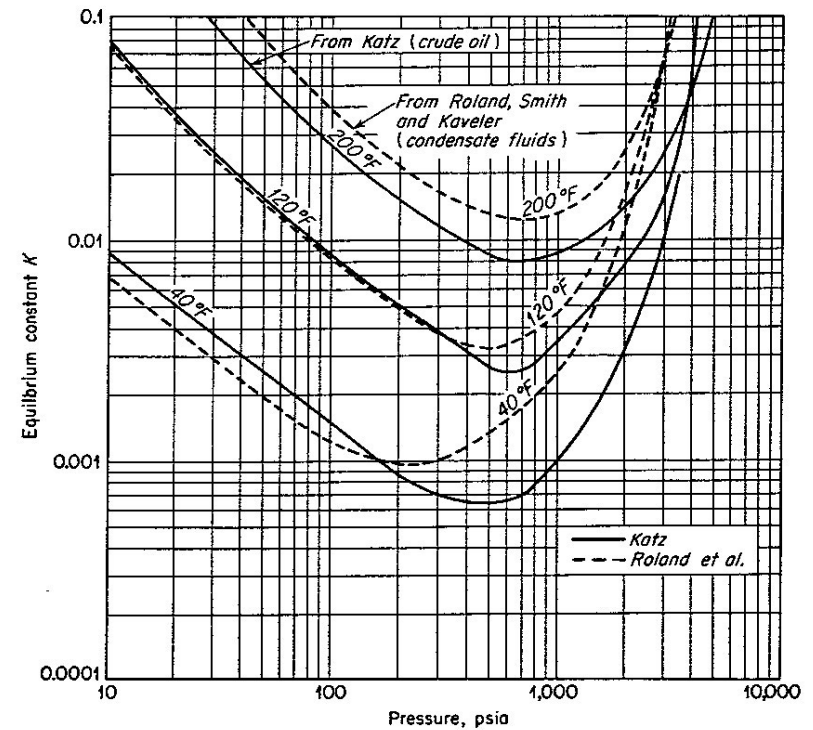


FIG. 4-81. Equilibrium ratios of heptanes-plus fraction. (Adapted from Katz and Hachmuth⁴¹ and Roland et al.⁴²)

The primary use of dew-point or bubble-point pressure calculations in petroleum-production engineering work is in connection with adjusting equilibrium ratios to observed bubble-point or dew-point pressures.

The calculation of phase equilibrium within the two-phase region involves the selection of equilibrium constants for the conditions of pressure and temperature under consideration and the solution of the relations expressed in Eqs. (4-77) through (4-80). Essentially two types of problems are of interest: the equilibrium between phases at reservoir conditions and at separator conditions. The calculative methods for these cases are identical except in the final volume calculation of the liquid. However, the

selection of appropriate equilibrium ratios for reservoir conditions is more difficult. The calculation of such equilibria based on adjusted equilibrium ratios will be discussed in Chap. 5.

For separator calculations at pressures of 500 psi or less, the 5,000-psi convergence pressure charts of the NGAA can be used for most naturally occurring hydrocarbon mixtures. The equilibrium ratio of the heptanes-plus fraction can be obtained from Fig. 4-81 or estimated as being equivalent to octane or nonane from the NGAA charts.

The procedure of solution is as follows:

1. Select a set of published equilibrium-ratio data which applies to the system of interest.
2. From the selected equilibrium-ratio data determine the value of K_i for each component at the desired pressure and temperature.
3. Assume a value of V , thus fixing the other value by the equation

$$L + V = 1 \quad (4-74)$$

4. Solve the following equations:

$$\sum_{i=1}^m x_i = \sum_{i=1}^m \frac{z_i}{L + VK_i} = 1 \quad (4-77)$$

5. If $\sum_{i=1}^m x_i = 1$, the problem is solved and the value assumed for V or L is the correct value. The values calculated for x_i 's represent the analysis of the liquid. If $\sum_{i=1}^m x_i \neq 1$, then the assumed value of V or L is not correct and it is necessary to assume another value of V or L and repeat step 4. This particular procedure is well suited to slide-rule calculations. It is perhaps a little slower than other procedures, but the point of convergence can be defined with fewer significant figures.

A variation of the calculating procedure which is particularly advantageous when using a desk calculator is as follows:

- 4a. Solve the equation

$$\sum_{i=1}^m \frac{z_i}{(L/V_a) + K_i} = (V)_{calc} \quad (4-78)$$

- 5a. If the calculated value $(V)_{calc}$ is equal to the assumed value V_a , the problem is solved and the composition of the liquid can be determined from

$$x_i = \frac{1}{V} \frac{z_i}{(L/V) + K_i} \quad (4-76)$$

and that of the vapor from

$$y_i = K_i x_i \quad \text{or} \quad y_i = \frac{z_i - x_i L}{V}$$

6a. If the calculated value $(V)_{calc}$ is not equal to the assumed value V_a , a new assumption for V must be made and the calculation repeated.

7a. By several successive trials V can be calculated to an accuracy of about 0.0001.

The procedure outlined is in most convenient form for solution with a desk calculator which has a cumulative multiplication feature. Successive trials can be made with a minimum transcription of numbers from the calculator to a table. A plot of the calculated V as a function of the assumed V is of aid in making estimates for later trials. The number of trials required depends on the experience of the engineer in estimating values of V .

The above relations apply when liquid is expected to be the major mole fraction. Although they are valid even though the vapor is the major fraction, it is desirable to use in that case either Eq. (4-79) or (4-80). This results in determining the composition of the major phase with greater accuracy.

The solution of equilibrium problems on high-speed computers has stimulated greater interest in flash calculations, particularly in solving reservoir equilibrium problems and multistage separator problems.

Rachford⁴⁴ has reported on a procedure which utilizes a form of the equilibrium relation developed as follows:

$$z_i = Lx_i + VK_i x_i \quad (4-83)$$

and

$$z_i = \frac{Ly_i}{K_i} + Vy_i$$

Since

$$L = 1 - V$$

$$x_i = \frac{z_i}{(K_i - 1)V + 1}$$

$$y_i = \frac{K_i z_i}{(K_i - 1)V + 1}$$

and

$$\sum x_i = \sum y_i = 1$$

Therefore,

$$\sum_{i=1}^m (y_i - x_i) = \sum_{i=1}^m \frac{(K_i - 1)z_i}{(K_i - 1)V + 1} = f(V, K_i, z_i) = 0 \quad (4-84)$$

must be assumed, and trial calculations performed. The correct value of V makes the function $f(V, K_i, z_i) = 0$. For other than the correct value

the graph of the function has the form shown in Fig. 4-82. The function has the unique property of yielding negative values if V (assumed) is too great and positive values if V (assumed) is too small. Trials can be made by always assuming $V = 0.5$ for the initial trial, then successively moving to higher or lower values in a systematic fashion by successive halving of the steps in the assumed values. This procedure yields rapid convergence on the correct value.

Another procedure proposed by Holland and Davison⁴⁵ uses the following form of the equilibrium relation:

$$g(L, K_i, z_i) = \sum_{i=1}^m \frac{z_i}{1 - L(1 - 1/K_i)} - 1 = 0 \quad (4-85)$$

In this relation L is assumed and the function is calculated. If the function is not equal to zero with the desired degree of accuracy, a new value

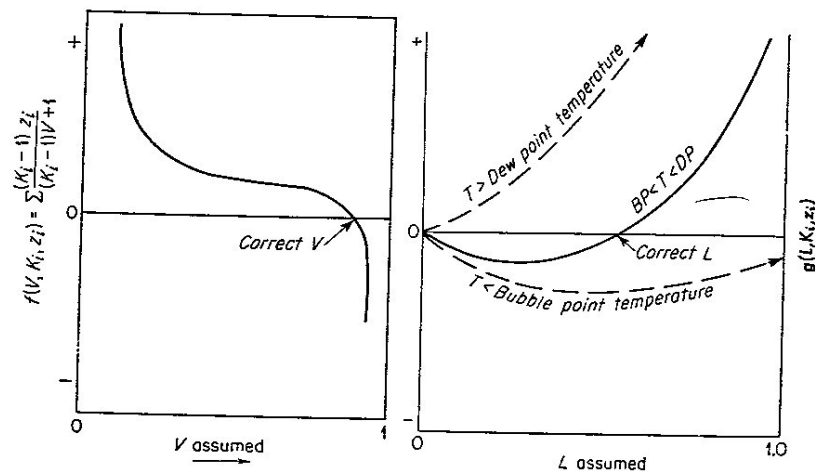


Fig. 4-82. Graph of Rachford's equilibrium function. (From Rachford and Rice.⁴⁴)

of L is assumed and the calculation repeated. A graph of the function is shown in Fig. 4-83. If, at the pressure selected, the selected temperature is greater than the dew-point temperature of the mixture, no solution exists except the trivial solution $L = 0$. Also if the selected temperature is less than the bubble-point temperature, the trivial solution $L = 0$ is the only solution obtained. The dashed curves represent solutions of such situations. The solid curve represents the function if the pressure and temperature conditions are truly in the two-phase region. In applying the method, a value of L very close to 1 is first assumed. The function is

evaluated. If the function is not equal to zero within the desired accuracy, a new L is assumed as follows:

$$L_j = L_{j-1} - \frac{g(L, K_i, z_i)_{j-1}}{g'(L, K_i, z_i)_{j-1}} \quad (4-86)$$

where $g'(L, K_i, z_i)_{j-1}$ is the first derivative of $g(L, K_i, z_i)_{j-1}$ and is equal to

$$g'(L, K_i, z_i) = \sum_{i=1}^m \frac{(1 - 1/K_i)z_i}{[1 - L(1 - 1/K_i)]^2} \quad (4-87)$$

L_j is the new assumed value of L , and L_{j-1} is the previous assumed value of L .

Separator Problems

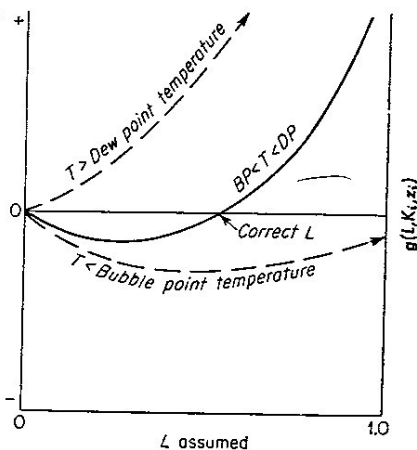
The application of equilibrium calculations to separator problems requires a brief discussion of the separation process. In a separator, a stream of fluid, referred to as the feed, is brought to equilibrium at the separator temperature and pressure. The pressure of the separator is subject to direct control by means of pressure-regulating devices. The temperature is usually determined by the temperature of the fluid entering the separator and the prevailing atmospheric temperature. Thus, the temperature of an oil-field separator may vary from a low at night to a high during the day. Seasonal variations also occur. In some instances separator temperatures are controlled by heating or by refrigeration.

The vapor and liquid are removed from contact on leaving the separator. Several separators may be operated in series, each receiving the liquid phase from the separator operating at the next higher pressure. Each condition of pressure and temperature at which vapor and liquid are separated is called a stage of separation. Examples of two- and three-stage separation processes are shown in Fig. 4-84. It can be noted that a process using one separator and a stock tank is a two-stage process unless the conditions of pressure and temperature of the two are identical.

Separator calculations are performed to determine the composition of the products, the oil volume factor, and the volume of gas released per barrel of oil and to determine optimum separator conditions for the particular conditions existing in a field.

Example 4-15 presents a sample calculation of two-stage separation applied to the reservoir fluid of Table 4-13. The equilibrium ratios were taken from Figs. 4-65 through 4-72 and 4-81. Two calculating procedures are presented in detail in the example. The calculations are illustrated in suitable form for the application of desk calculators. Graphs for estimating the value of V to be used in additional trials after the first two trials are illustrated in Figs. 4-85 and 4-86 for methods A and B, respectively. In method A, V (assumed) is plotted for each calculated Σz_i . The correct

Fig. 4-83. Graph of Holland's equilibrium function. (From Holland and Davison.⁴⁵)



value of V corresponds to $\sum x_i = 1$. Therefore, the extrapolation of a curve through the calculated points to the line $\sum x_i = 1$ yields a corrected estimate of V . For method B , V (assumed) is plotted for each V (calculated). The correct solution should lie on a 45° line (slope of 1) which passes

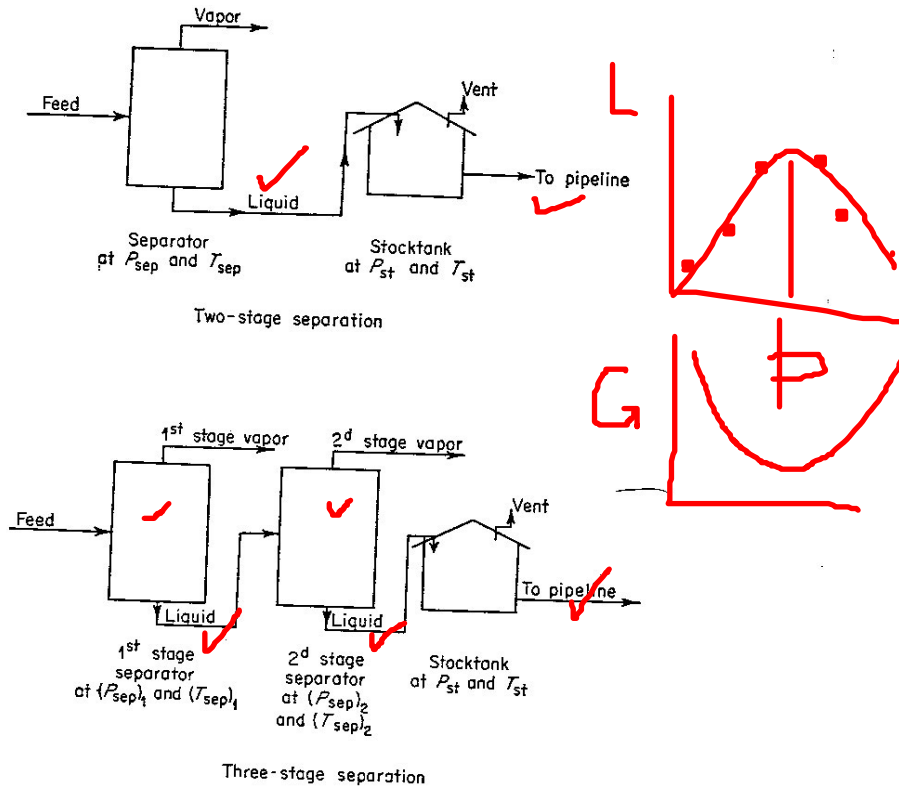


Fig. 4-84. Schematic drawing of separation processes.

through the origin. The intersection of the 45° and a curve through the calculated points yields the new estimate for V .

The calculation procedure by method A yields directly the values of x_i , the mole fraction of a component in the liquid. The mole fractions of the components in the vapor are calculated from the definition of the equilibrium ratio.

$$y_i = K_i x_i$$

Frequently, a satisfactory solution for the values of L , V , and x_i will be

obtained from which the calculated values of y_i do not sum to 1. This is because of minor discrepancies in the K values, and the error can be distributed uniformly over the composition of the vapor to yield corrected

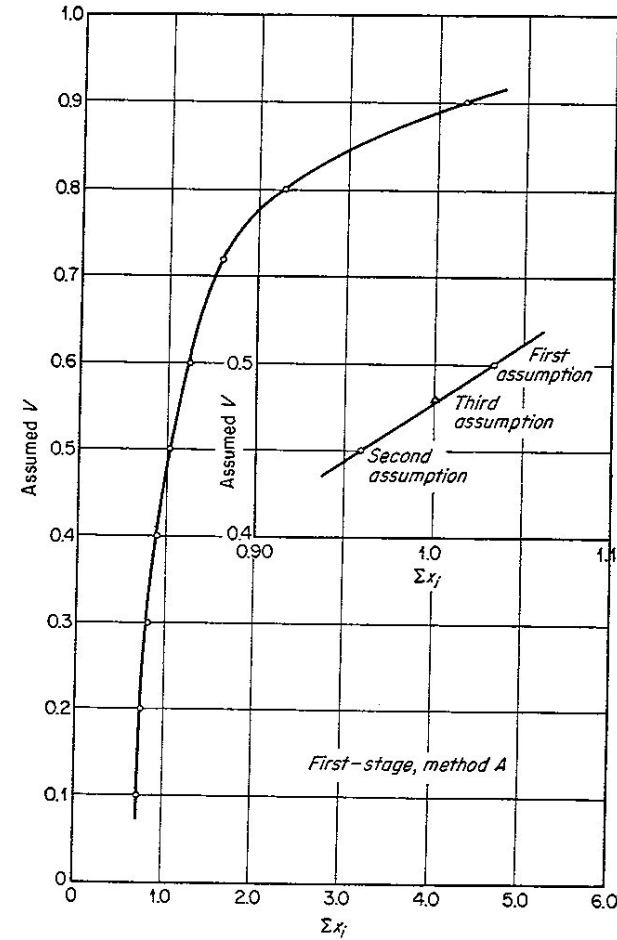


Fig. 4-85. Solution of equilibrium ratio equations, Eq. (4-77), for various assumed gas fractions.

y_i values, or they can be calculated by $y_i = (z_i - x_i L)/V$. For this reason, if it is expected that the vapor phase composition is critical, the alternative

relation $\sum \frac{z_i}{(L/K_i) + V} = 1$, based on $\sum y_i$, should be used.

In method B, after a satisfactory V is obtained, the x_i are calculated from

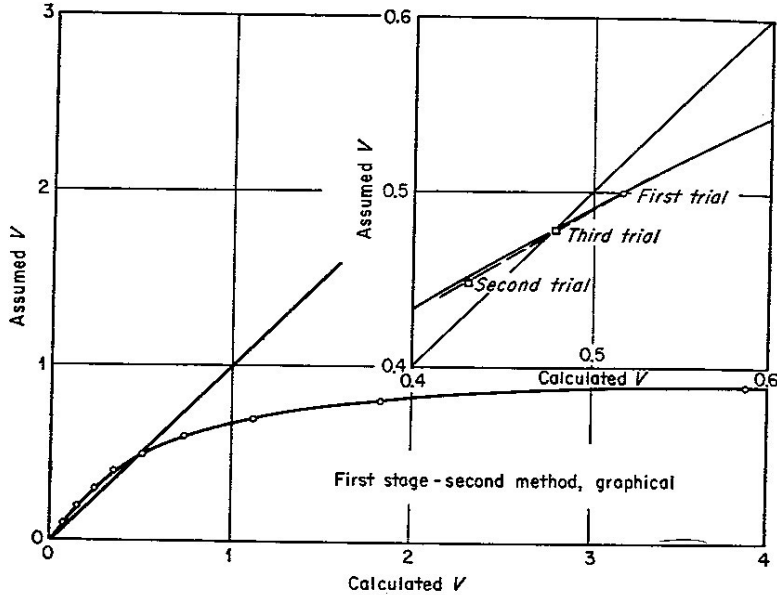


Fig. 4-86. Solution of equilibrium ratio equations, Eq. (4-78), for various assumed gas fractions.

$$x_i = \frac{1}{V(L/V) + K_i} z_i \quad (4-76)$$

and y_i from

$$y_i = K_i x_i$$

Once again it may be necessary to adjust the values of y_i. If the vapor phase composition is desired, the relation, based on $\sum y_i$,

$$\sum_{i=1}^m \frac{z_i}{(L/K_i V) + 1} = V \quad (4-80)$$

should be used.

The equation $z_i = x_i L + y_i V$ can be used to calculate the composition of the second phase, and no correction will be necessary. This method is longer, but if the problem merits the additional accuracy, it is desirable.

Example 4-15. Calculation of Two-stage Separation Problem.

**CALCULATION FOR FIRST-STAGE SEPARATION (35 PSIA, 40°F)
METHOD A [Eq. (4-77)]**

| Component | Reservoir fluid analysis z _i | K _i at 35 psia and 40°F ^a | First assumption: L = 0.5, V = 0.5 | | | Second assumption: L = 0.65, V = 0.46 | | | Third assumption: L = 0.52060, V = 0.47804 | | | Y _i = x _i K _i |
|-----------------|---|---|------------------------------------|---------------------|------------------------------|---------------------------------------|---------------------|------------------------------|--|---------------------|------------------------------|--|
| | | | VK _i | L + VK _i | $\frac{z_i}{L + VK_i} = x_i$ | VK _i | L + VK _i | $\frac{z_i}{L + VK_i} = x_i$ | VK _i | L + VK _i | $\frac{z_i}{L + VK_i} = x_i$ | |
| C ₁ | 0.3396 | 61.0000 | 30.5000 | 31.0000 | 0.0109 | 27.4500 | 0.0121 | 29.2214 | 0.0114 | 0.0965 | | |
| C ₂ | 0.0546 | 9.000 | 4.5000 | 5.0000 | 0.0129 | 4.0500 | 0.0150 | 4.3113 | 0.0133 | 0.1203 | | |
| C ₃ | 0.0987 | 2.2000 | 1.0000 | 1.6000 | 0.0616 | 0.9900 | 0.0641 | 1.0639 | 0.0627 | 0.1379 | | |
| C ₄ | 0.0434 | 0.0100 ^b | 0.3050 | 0.8050 | 0.0539 | 0.2745 | 0.0526 | 0.2922 | 0.8131 | 0.0326 | | |
| C ₅ | 0.0320 | 0.1570 ^b | 0.0786 | 0.5785 | 0.0553 | 0.0706 | 0.0510 | 0.0752 | 0.6061 | 0.0084 | | |
| C ₆ | 0.0300 | 0.0350 | 0.0173 | 0.5175 | 0.0579 | 0.0157 | 0.5057 | 0.0107 | 0.5377 | 0.0019 | | |
| C ₇₊ | 0.3917 | 0.0032 | 0.0016 | 0.5016 | 0.7809 | 0.0014 | 0.7103 | 0.0015 | 0.5224 | 0.0024 | | |
| Total | 1.0000 | | | | 1.0330 | | 0.9590 | | 1.0000 | | | |

^a From Figs. 4-65 through 4-72 and 4-81.
^b Average of iso and normal.

Equations used:

$$x_i = \frac{z_i}{L + VK_i} \quad y_i = K_i x_i \quad \sum_{i=1}^m y_i = 1 \quad y_i = \frac{y_i}{\sum_{i=1}^m y_i}$$

Final answer:
0.52060 mole liquid
0.47804 mole vapor

~~CALCULATION FOR FIRST-STAGE SEPARATION (35 PSIA, 40°F)
METHOD B [Eq. (4-75)]~~

| Component | Reservoir fluid analysis z_i | K_i at 35 psia and 40°F ^a | $L = 0.5, V = 0.5$ | | $L = 0.55, V = 0.45$ | | $L = 0.52096, V = 0.47904$ | | z_i | $y_i = K_i z_i$ |
|-----------------|--------------------------------|--|---------------------|-----------------------|----------------------|-----------------------|----------------------------|-----------------------|--------|-----------------|
| | | | $\frac{L}{V} + K_i$ | $\frac{z}{L/V + K_i}$ | $\frac{L}{V} + K_i$ | $\frac{z}{L/V + K_i}$ | $\frac{L}{V} + K_i$ | $\frac{z}{L/V + K_i}$ | | |
| C ₁ | 0.3396 | 61.0000 | 62.0000 | 0.0055 | 62.2222 | 0.0054 | 62.0875 | 0.0055 | 0.0114 | 0.6965 |
| C ₂ | 0.0646 | 9.0000 | 10.0000 | 0.0065 | 10.2222 | 0.0063 | 10.0875 | 0.0064 | 0.0133 | 0.1203 |
| C ₃ | 0.0987 | 2.2000 | 3.2000 | 0.0308 | 3.4222 | 0.0288 | 3.2875 | 0.0300 | 0.0627 | 0.1379 |
| C ₄ | 0.0434 | 0.6100 ^b | 1.6100 | 0.0269 | 1.8322 | 0.0237 | 1.6975 | 0.0255 | 0.0533 | 0.0326 |
| C ₅ | 0.0320 | 0.1570 ^b | 1.1570 | 0.0276 | 1.3792 | 0.0232 | 1.2445 | 0.0257 | 0.0537 | 0.0084 |
| C ₆ | 0.0300 | 0.0350 | 1.0350 | 0.0290 | 1.2572 | 0.0239 | 1.1225 | 0.0268 | 0.0559 | 0.0019 |
| C ₇₊ | 0.3917 | 0.0032 | 1.0032 | 0.3905 | 1.2254 | 0.3196 | 1.0907 | 0.3591 | 0.7497 | 0.0024 |
| Total | | | | $V = 0.5108$ | | $V = 0.4309$ | | $V = 0.4790$ | 1.0000 | 1.0000 |

^a From Figs. 4-65 through 4-72 and 4-81.

^b Average of iso and normal.

Equations used:

$$\sum_{i=1}^m \frac{z_i}{L/V + K_i} = V \quad z_i = \frac{1}{V} \frac{z_i}{L/V + K_i}$$

$$y_i = K_i z_i \quad y_{7+} = \frac{y_i}{\sum_{i=1}^m y_i}$$

Final answer:

0.52096 mole liquid
0.47904 mole vapor

CALCULATION FOR SECOND STAGE OF SEPARATION (15 PSIA, 40°F)
METHOD A

| Component | Reservoir fluid analysis z_i | K_i at 15 psia and 40°F ^a | First assumption: $L = 0.95, V = 0.05$ | | | Second assumption: $L = 0.9789, V = 0.0211$ | | | z_i | $y_i = K_i z_i$ | y_i corrected |
|-----------------|--------------------------------|--|---|------------|----------------------------|--|------------|----------------------------|--------|-----------------|-----------------|
| | | | VK_i | $L + VK_i$ | $\frac{z}{L + VK_i} = x_i$ | VK_i | $L + VK_i$ | $\frac{z}{L + VK_i} = x_i$ | | | |
| C ₁ | 0.0114 | 145.0000 | 7.2500 | 8.2000 | 0.0013 | 3.0595 | 4.0384 | 0.0028 | 0.4103 | 0.4109 | |
| C ₂ | 0.0133 | 20.5000 | 1.0250 | 1.9750 | 0.0067 | 0.4325 | 1.4114 | 0.0095 | 0.1941 | 0.1944 | |
| C ₃ | 0.0627 | 5.1000 | 0.2550 | 1.2050 | 0.0520 | 0.1076 | 1.0805 | 0.0577 | 0.2942 | 0.2945 | |
| C ₄ | 0.0534 | 1.4000 ^b | 0.0700 | 1.0200 | 0.0528 | 0.0205 | 1.0084 | 0.0529 | 0.0741 | 0.0742 | |
| C ₅ | 0.0637 | 0.3750 ^b | 0.0178 | 0.9678 | 0.0554 | 0.0075 | 0.9864 | 0.0544 | 0.0194 | 0.0194 | |
| C ₆ | 0.0558 | 0.0750 | 0.0037 | 0.9537 | 0.0584 | 0.0016 | 0.9805 | 0.0569 | 0.0043 | 0.0043 | |
| C ₇₊ | 0.7497 | 0.0030 | 0.0001 | 0.9501 | 0.7890 | 0.0006 | 0.9789 | 0.7658 | 0.0023 | 0.0023 | |
| Total | 1.0000 | | | | 1.0154 | | | 1.0000 | 0.9987 | 1.0000 | |

^a From Figs. 4-65 through 4-72 and 4-81.

^b Average of iso and normal.

Final answer:

0.9789 mole liquid
0.0211 mole vapor

In addition to the compositions and the mole fractions of the mixture which exist in the liquid and vapor phases, three additional parameters are of interest to the engineer. These are the API gravity of the stock-tank liquid, the gas-oil ratio, and the oil-formation volume factor. The API gravity can be calculated from the density of the stock-tank liquid determined from the composition by the procedure of Standing and Katz²⁷ discussed earlier in this chapter.

The gas-oil ratio is defined as the volume of gas in standard cubic feet per barrel of stock-tank oil. Gas is removed from each stage of the separation process so that the gas-oil ratio can be calculated for each stage or combination of stages. Total gas-oil ratio refers to the sum of the gas volumes in standard cubic feet from all stages divided by the volume in barrels of stock-tank oil.

In equilibrium (flash) calculations it is customary to solve the equilibrium relations for each stage on the basis of 1 mole of feed to that stage. Thus, if n_1 moles of feed enter the first stage, the moles of liquid entering the second stage

$$\begin{aligned} n_2 &= L_1 n_1 \\ n_3 &= L_2 n_2 = L_2 L_1 n_1 \end{aligned}$$

and where n is the moles of feed and the subscripts refer to the stage. If the third stage is the stock tank, then

$$n_{s,t} = L_3 n_3 = L_3 L_2 L_1 n_1$$

where $n_{s,t}$ is the moles of liquid remaining in the stock tank for n_1 moles into the first separator. In general terms,

$$n_{s,t} = n_1 \prod_{i=1}^m L_i \quad (4-88)$$

where m = number of stages

L_i = mole fraction of liquid off i th stage

n_1 = moles of feed to first stage

If $n_1 = 1$, then

$$\bar{n}_{s,t} = \prod_{i=1}^m L_i \quad (4-89)$$

and $\bar{n}_{s,t}$ is the mole fraction of stock-tank oil in the feed.

In a similar manner the number of moles of gas can be evaluated. Let $n_{g,i}$ be the moles of gas off stage i . Then

$$n_{g1} = V_1 n_1$$

The moles of gas off the second stage

$$n_{g2} = V_2 n_2 = V_2 L_1 n_1$$

and from the third stage,

$$n_{g3} = V_3 n_3 = V_3 L_2 L_1 n_1$$

In general terms the total gas off all stages

$$n_{gT} = \sum_{i=1}^m n_{g,i} = n_1 \sum_{i=1}^m V_i \prod_{j=1}^{i-1} L_j = n_1 - n_{s,t} \quad (4-90)$$

If $n_1 = 1$, then

$$\bar{n}_{gT} = \sum_{i=1}^m V_i \prod_{j=1}^{i-1} L_j \quad (4-91)$$

where \bar{n}_{gT} is the mole fraction of total gas in the feed, such that $n_1 = \bar{n}_{s,t} + \bar{n}_{gT}$. The number of moles of gas can be readily converted to standard cubic feet by multiplying the number of moles by the molal volume V_m from Table 4-5 for the desired standard conditions. Thus the total gas volume per mole of feed = $\bar{n}_{gT} V_m$.

The volume of stock-tank oil per mole of feed can be calculated from the density and molecular weight of the stock-tank oil as follows:

$$(V_{s,t})_m = \frac{\bar{n}_{s,t} M_{s,t}}{\rho_{s,t}} \quad (4-92)$$

where $(V_{s,t})_m$ = volume of stock-tank oil per mole of feed, bbl

$M_{s,t}$ = molecular weight of stock-tank oil

$\bar{n}_{s,t}$ = moles of stock-tank oil per mole of feed

$\rho_{s,t}$ = density of stock-tank oil at 60°F and atmospheric pressure, lb/bbl

The total gas-oil ratio is given by

$$R_T = \frac{\bar{n}_{gT} V_m}{(V_{s,t})_m} = \frac{\bar{n}_{gT} V_m \rho_{s,t}}{\bar{n}_{s,t} M_{s,t}} \quad (4-93)$$

where R_T is the total gas-oil ratio.

If it is known from other data that the feed to the first-stage separator exists as a single-phase liquid at its point of entry into the production stream, then an oil-formation volume factor B_o can be calculated from the data obtained. The analysis of the feed can then be treated as the reservoir fluid, and its density calculated by the method of Standing and Katz.

Let ρ_{res} be the density of the feed in pounds per barrel. Then

$$(V_{res})_m = \frac{M_{res}}{\rho_{res}} \quad (4-94)$$

where $(V_{res})_m$ is the volume occupied by 1 mole of feed at reservoir conditions and M_{res} is the molecular weight of the feed. By definition, the oil-formation volume factor

$$B_o = \frac{(V_{res})_m}{(V_{st})_m} = \frac{M_{res} \rho_{st}}{\rho_{res} M_{st} \bar{v}_{st}} \quad (4-95)$$

The calculation of API gravity, total gas-oil ratio, and oil formation volume factor is illustrated in Example 4-16.

The application of equilibrium calculations to the determination of the optimum first-stage separator pressure in a two-stage system (one separator and the stock tank) is illustrated by the data of Table 4-17 and the graphs in Fig. 4-87. The optimum pressure is defined as that pressure at which

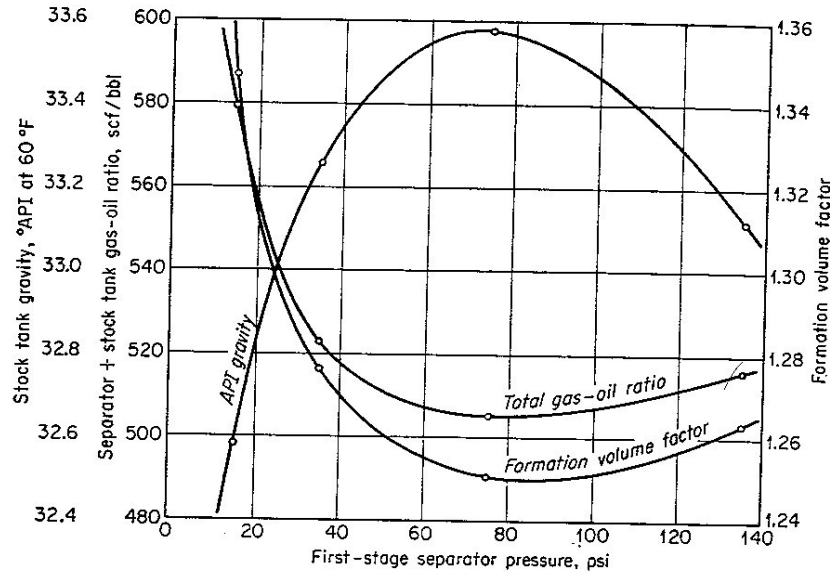


Fig. 4-87. Effect of separator pressure in a two-stage separation process.

the maximum liquid volume is accumulated in the stock tank per volume of reservoir fluid produced. This pressure corresponds to a maximum in the API gravity and a minimum in the gas-oil ratio and oil formation volume factor. The optimum first-stage separator pressure for the system evaluated in Table 4-17 and Fig. 4-87 is 75 psia. The effect of additional stages can be evaluated in a similar manner.

Equilibrium, or "flash," calculations can be used in many other applications. The applications of the methods to calculating the performance of reservoirs containing condensate fluids or volatile oils are becoming increasingly important. Many of these applications will be reviewed later in this text.

Example 4-16. Calculation of Stock-tank Gravity Separator and Stock-tank Gas-Oil Ratios and Formation Volume Factor.

| (1) Component | (2) Mole fraction in liquid phase at 15 psia, 40°F ^a | (3) Mole wt, lb/mole ^b | (4) Relative wt, lb/mole (2) × (3) | (5) Liquid density at 60°F, 14.7 psia ^b | (6) Liquid volume cu ft/mole (4) ÷ (5) |
|------------------|---|---|---|--|--|
| C ₁ | 0.00283 | 16.042 | 0.045399 | | |
| C ₂ | 0.00947 | 30.068 | 0.284744 | | |
| C ₃ | 0.05768 | 44.094 | 2.543342 | 31.64 | 0.0803837 |
| C ₄ | 0.05292 | 58.120 | 3.075710 | 35.71 ^d | 0.0861302 |
| C ₅ | 0.05441 | 72.146 | 3.925464 | 39.08 ^d | 0.1004468 |
| C ₆ | 0.05690 | 86.172 | 4.903187 | 41.36 | 0.1185490 |
| C ₇₊ | 0.76579 | 263 ^c | 201.402770 | 55.28 | 3.6433207 |
| Total | 1.00000 | | 216.180616 | | 4.0108304 |

^a From Example 4-15.

^b From Table 4-4.

^c From Table 4-13.

^d Average of iso and normal.

$$\text{Density of } C_{3+} = \frac{\text{wt of } C_{3+}}{\text{vol of } C_{3+}} = \frac{215.850}{4.011} = 53.817 \text{ lb/cu ft}$$

$$\text{Wt } \% C_2 \text{ in } C_{2+} = \frac{0.285}{216.135217} \times 100 = 0.132$$

$$\text{Wt } \% C_1 \text{ in } C_{1+} = \frac{0.046}{216.181} \times 100 = 0.021$$

Density of system including C₁ and C₂ = 53.5 lb/cu ft at 60°F and 14.7 psia (from Fig. 4-50)

Pressure correction from 14.7 to 15 psia is negligible. No temperature correction required.

Density = 53.5 lb/cu ft at 60°F and 15 psia

Specific gravity = 0.8588

API gravity = 33.26°

$$\begin{aligned} \text{Separator GOR (gas-oil ratio)} &= \frac{(380.69 V_i) \rho_{st}}{M_{st} L_1 L_2} \\ &= \frac{380.69(.47904)(53.5)(5.61)}{216.181(.52096)(.9789)} \\ &= 496.47 \text{ cu ft/bbl} \end{aligned}$$

$$\begin{aligned} \text{Stock-tank (ST) GOR} &= \frac{380.69V_{20.1}(5.61)}{\text{mol wt ST liq.} \times L_2} \\ &= \frac{380.69(0.0211)(53.5)(5.61)}{216.181(.9789)} \\ &= 11.35 \text{ cu ft/bbl.} \end{aligned}$$

$$\begin{aligned} \text{Total GOR} &= \text{separator GOR} + \text{stock-tank GOR} \\ &= 496.47 + 11.35 = 507.82 \text{ cu ft/bbl} \end{aligned}$$

Sample calculation of formation volume factor:

$$B_o = \frac{\text{bbl res liq/mole res liq}}{\text{ST bbl/mole res liq}} \quad (4-95)$$

$$\begin{aligned} \text{Reservoir bbl/mole res liquid} &= \frac{M_{\text{res}}}{P_{\text{res}}} \\ &= \frac{122.1748}{46.6(5.61)} = 0.467 \end{aligned}$$

$$\begin{aligned} \text{Stock-tank vol/mole stock-tank oil} &= \frac{216.181}{53.5(5.61)} \\ &= 0.72027792 \text{ bbl/mole} \end{aligned}$$

$$\begin{aligned} \text{Stock-tank bbl/mole res fluid} &= \text{bbl/mole ST liq} \times L_1 L_2 \\ &= 0.720(0.52096)(0.9789) \\ &= 0.367 \end{aligned}$$

$$B_o = \frac{0.467}{0.367} = 1.2723$$

TABLE 4-17. EFFECT OF SEPARATOR PRESSURES ON FLUID PROPERTIES

| Separator pressure, psia ^a | | Gas-oil ratio, cu ft/bbl | | Total | Stock-tank oil gravity, °API | B _o , oil formation volume factor ^b |
|---------------------------------------|-----------------------|--------------------------|------------|-------|------------------------------|---|
| 1st stage | 2d stage (stock tank) | Separator | Stock tank | | | |
| 15 | 15 | 587.2 | | 582.2 | 32.58 | 1.340 |
| 35 | 15 | 507.7 | 8.5 | 516.2 | 33.26 | 1.283 |
| 75 | 15 | 450.4 | 39.9 | 490.3 | 33.58 | 1.265 |
| 135 | 15 | 368.1 | 135.7 | 503.8 | 33.12 | 1.276 |

^a Separator and stock-tank temperature, 40°F.

^b Reservoir pressure, 3,614 psia; reservoir temperature, 200°F.

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