

This procedure for using E_{MV} works well for calculations upward in a tower but is more difficult for calculations downward. An expedient for the case where E_{MV} is the same on all stages is shown in Fig. 5-24. A new curve, shown dashed in Fig. 5-24, is drawn in and is so located that it always lies a fraction E_{MV} of the vertical distance to the equilibrium curve from the appropriate operating lines. Actual stages can then be stepped off, as shown, using this curve rather than the equilibrium curve. The dashed curve in Fig. 5-24 corresponds to $E_{MV} = 0.67$, including the reboiler.

Means of predicting and correlating stage efficiencies are covered in Chap. 12.

OTHER PROBLEMS

The solution of a typical design problem has been shown at some length because these problems constitute a large proportion of those encountered and because they can be solved in a straightforward fashion on the McCabe-Thiele diagram. Any other problem which might be encountered in binary distillation also can be solved on the diagram but not necessarily in a straightforward way. Trial-and-error procedures are often required, and the exact way to approach the solution on the diagram must be thought out. In some cases an algebraic approach is as efficient as the graphical approach or more so.

Example 5-1 illustrates the solution of a problem concerning the operation of an existing still.

Example 5-1 A binary mixture is to be separated in a column which contains five *equilibrium* rectifying stages plus a reboiler. The feed is a saturated liquid at column pressure and is introduced into the reboiler. A total condenser is used, and the reflux will be returned to the column at its saturation temperature. The reflux rate is 0.5 mol per unit time. The feed rate is 1 mol per unit time, and $z_{A,F} = 0.5$. The mole fraction of component A in the top product is to be 0.90. Assume that the relative volatility is constant at $\alpha_{AB} = 2$ for the column pressure and the temperature spread of the column. Also assume constant molal overflow. Calculate the mole fraction of component A in the bottom product and calculate the amounts of top and bottom products.

SOLUTION From the foregoing description, the column is as shown in Fig. 5-25. For such a column the number of variables to be set in any problem description is counted as

| | |
|---------------------------|-------|
| $Fz_{A,F}$ and $Fz_{B,F}$ | n |
| h_F | Q_C |
| Pressure | Q_R |
| Reflux temperature | |

These are the variables which can be set structurally or by external manipulation during operation. In the actual problem description the following variables are set:

| | |
|------------------------------|--|
| $Fz_{A,F}$ and $Fz_{B,F}$ | $F = 1$ mol, $z_{A,F} = 0.5$, $z_{B,F} = 0.5$ |
| h_F | Feed is saturated liquid $L_F = F$ |
| Pressure | $\alpha_{AB} = 2$ |
| Reflux temperature | Reflux is saturated |
| n | $n = 5$ |
| r (replacing Q_C) | $r = 0.5$ mol |
| $x_{A,d}$ (replacing Q_R) | $x_{A,d} = 0.90$ |

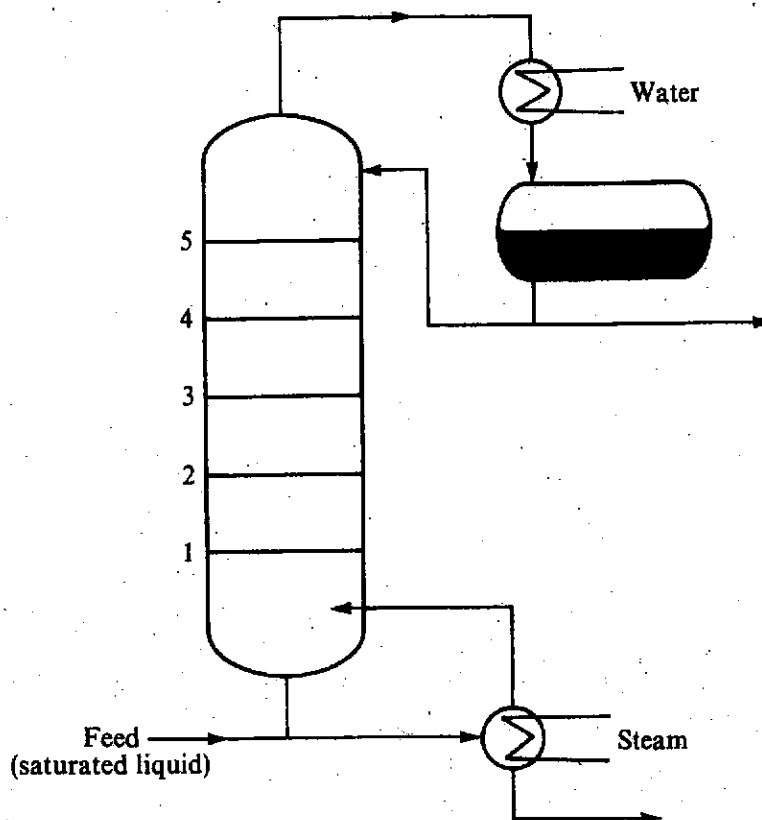


Figure 5-25 Column for Example 5-1.

The equilibrium curve can be drawn on the xy diagram from Eq. (1-12)

$$y_A = \frac{2x_A}{1 + x_A} \quad (5-34)$$

For various values of x_A between 0 and 1, the corresponding values of y_A in equilibrium are calculated and plotted on the xy diagram, as shown in Fig. 5-26.

It can be assumed that a good estimate of the liquid flow in the rectifying section will be $L = r$, since the reflux is saturated. When $x_{A,d}$ is known but d is not known, the rectifying-section operating line cannot be plotted. The locus of operating-line intersections can be plotted and is a vertical line through $z_{A,r} = 0.5$.

Since all independent variables have been set, the total amounts of products and the bottoms product composition are dependent variables whose values are unknown. If a value of d is assumed, the corresponding values of b and $x_{A,b}$ can be calculated by overall mass balance. The assumed value of d will also yield a particular set of operating lines, and consequently five equilibrium stages plus the reboiler can be stepped off on the diagram from the top down. If the assumed value of d is correct, the last vertical leg on the diagram, representing $x_{A,b}$, will be at the same value of $x_{A,b}$ that was obtained from the overall mass balance. If not, another value of d must be assumed until the correct value is found. The solution procedure therefore involves assuming values for one dependent variable d , then calculating another dependent variable $x_{A,b}$ by two different routes until both routes give the same value of $x_{A,b}$. A typical solution follows.

Assume $d = 0.25$ This is done since the column is inefficient for stripping A out of the bottom product (why?) and hence should not make a large amount of the high-purity A top product.

$$x_{A,d}d = (0.25)(0.9) = 0.225 \quad V = L + d = r + d = 0.5 + 0.25 = 0.75$$

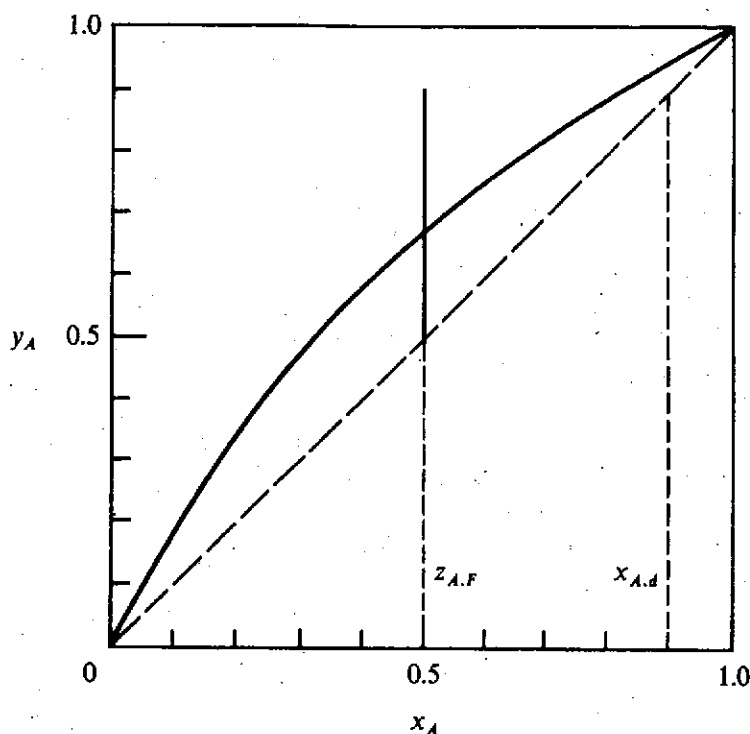


Figure 5-26 Initial operating diagram for Example 5-1.

The intercept of the operating line at $x_A = 0$ is

$$\frac{dx_{A,d}}{V} = \frac{0.225}{0.75} = 0.300$$

$$x_{A,b}b = Fx_{A,F} - x_{A,d}d = 0.5 - 0.225 = 0.275$$

$$b = F - d = 1 - 0.25 = 0.75 \quad x_{A,b} = \frac{0.275}{0.75} = 0.367$$

Figure 5-27 shows the operating line corresponding to $d = 0.25$ and the construction of steps corresponding to the stages. Note that each time a step is made to an operating line, it is the rectifying-section operating line which is appropriate. From the construction, the vertical leg from the reboiler is at $x_A = 0.47$; this is above the value of $x_{A,b} = 0.367$, which was calculated from the overall mass balance. Hence $d = 0.25$ is wrong. If the next guess of the amount of top product is lower, there will be more A in the bottom product and $x_{A,b}$ calculated from the overall mass balance will be higher on the diagram. Also the slope of the rectifying operating line will be closer to unity and the construction of stages will move further down the diagram. These are opposing effects; hence the calculation will converge readily, and the next assumed value for d should be less than 0.25.

Assume $d = 0.18$

$$x_{A,d}d = (0.18)(0.9) = 0.162 \quad V = 0.5 + 0.18 = 0.68$$

The intercept at $x_A = 0$ is

$$\frac{0.162}{0.68} = 0.238$$

$$x_{A,b}b = 0.5 - 0.162 = 0.338$$

$$b = 1 - 0.18 = 0.82 \quad x_{A,b} = \frac{0.338}{0.82} = 0.412$$

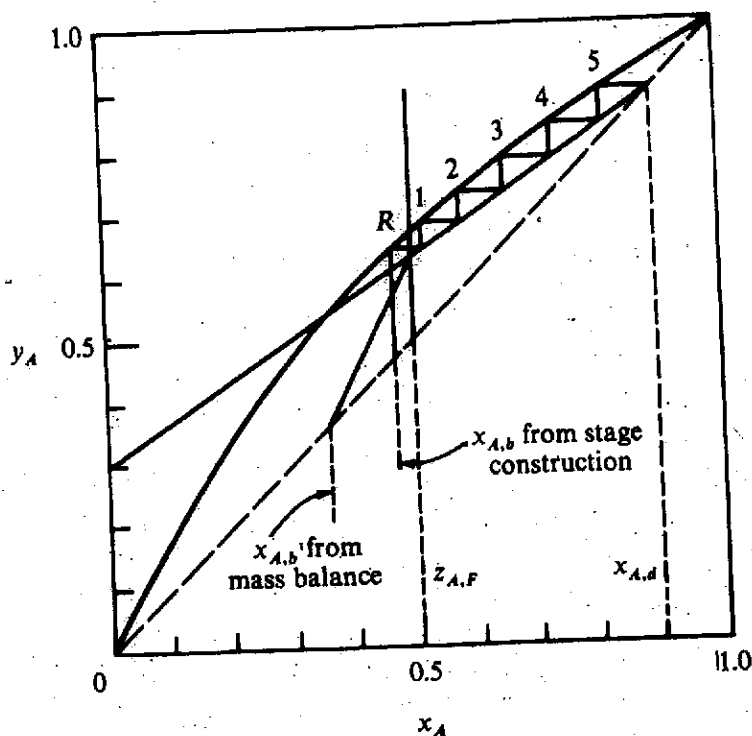


Figure 5-27 First-trial construction for Example 5-1, with $d = 0.25$.

Figure 5-28 shows the new operating lines and new construction. The vertical leg from the reboiler is at $x_A = 0.40$, which is now below the value of $x_{A,b} = 0.412$ obtained from the mass balance. Another assumption of d could be made between $d = 0.25$ and $d = 0.18$, but the inaccuracy in construction does not really justify it. Linear interpolation between the two values is probably the best way to arrive at the answer. On this basis

$$d = 0.18 + (0.25 - 0.18) \frac{0.012}{0.012 + 0.103} = 0.186$$

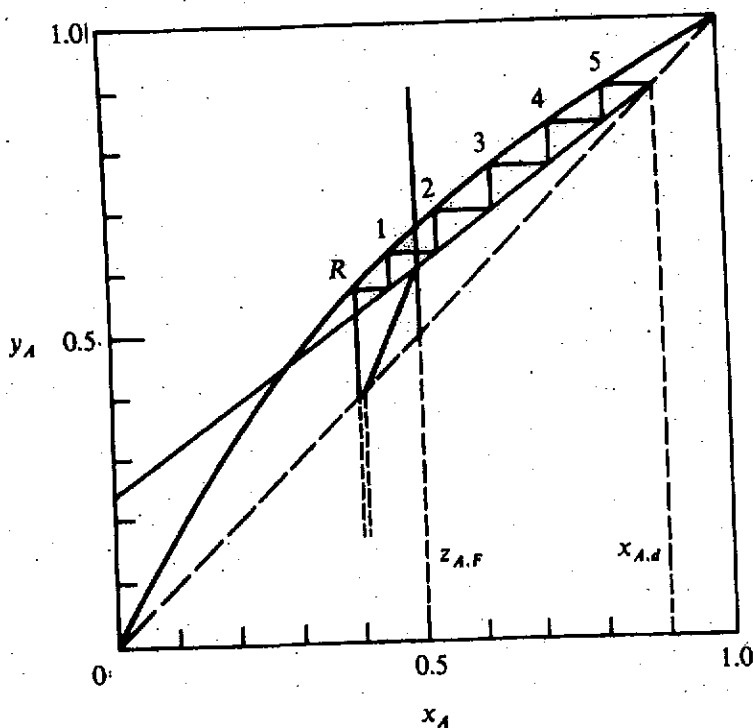


Figure 5-28 Second-trial construction for Example 5-1, with $d = 0.18$.

Thus the answers required are $d = 0.186$ mol, $b = 0.814$ mol, and $x_{A,b} = 0.408$. It might be noted that the results of the calculation show that a very small fraction of component A in the feed has been recovered as a relatively pure top product. The amount of recovery could be increased by using a feed point higher in the column. □

MULTISTAGE BATCH DISTILLATION

Multistage distillations also can be run on a batch basis. In the column shown in Fig. 5-29 an initial charge of liquid is fed to the still pot. The heating and cooling media are then turned on, and distillation proceeds, continually depleting the liquid in the still pot and building up overhead product in the distillate receiver. The operation is therefore the same as the simple Rayleigh distillation shown in Fig. 3-8 except for the presence of plates above the still pot and for the manufacture of reflux.

Batch stills require considerably more labor and attention than continuous columns. It is also necessary to shut down, drain, and clean the column in between charges, and this can result in a substantial loss of on-stream time. Consequently batch multistage distillation is most often employed when a product is to be manufactured only at certain isolated times and where a number of different mixtures can be handled at different times by the same column. Batch distillations are more common in smaller, multiproduct plants.

In a batch distillation the compositions at all points in the column are continually changing. As a result a steady-state analysis of the type employed for continuous distillation cannot be made of the column behavior. On each plate a mixing process is occurring such that

Input - output = accumulation

$$V_{p-1}y_{A,p-1} + L_{p+1}x_{A,p+1} - V_p y_{A,p} - L_p x_{A,p} = \frac{d}{dt}(Mx_{A,p}) \quad (5-35)$$

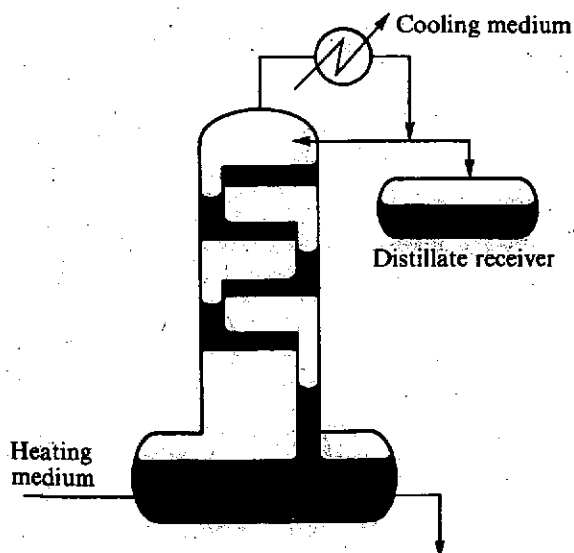


Figure 5-29 Batch multistage distillation.