Binary Distillation Design

Contents

These notes deal with the design of binary distillation columns, that is, distillation columns that separate a mixture of two components into two product streams, a *distillate* and a *bottoms product*. The distillate leaves at the top of the column and carries most of low-boiling, more volatile or *light* component, while the bottoms product carries most of the high-boiling, less volatile or *heavy* component. Distillation separates the components based on the difference in their volatilities. The

presentation will be restricted to the design of tray columns, that is, columns that carry out the separation by countercurrent contact of the liquid and vapor phases in trays or plates.
Today chemical engineers use process simulators (e.g., Aspen Plus, HySys, Pro II) to carry out the complex calculations involved in the design of a column. Distillation columns are some of the most expensive units to purchase, install, and operate in a process. So it is important for the engineer to understand how the design decisions made when simulating the column affect the size and cost of the column. Here we will use the McCabe-Thiele graphical procedure to acquire an understanding of the effect of the different design decisions have on the size of the column. Such an understanding is essential for you to become proficient in the use of the column simulators.

Next we look at the variables associated with the feed to the column and how they affect the design of the column.

1. Feed Variables

On a distillation column with one feed and two product streams there are $C + 6$ degrees of freedom, where C is the number of components in the feed. Of these, $C + 1$ are feed variables: the feed rate, $C - 1$ component mole fractions, and enthalpy, leaving five degrees of freedom for the design of the column. When a column has more than one feed stream each feed has $C + 1$ variables associated with it. In this sub-section we look at the feed variables and their effect on the column design.
Feed Rate. Although the feed rate has no effect on the number of stages required to do the separation, for a given design, the

vapor and liquid rates in the column are proportional to the feed rate. So the heat duties and sizes of the condenser and the reboiler are proportional to the feed rate and the column diameter is proportional to the square root of the feed rate as the column area, $\pi D^2/4$, is proportional to the vapor rate. When sequencing distillation columns for the separation of multicomponent mixtures one of the strategies is to arrange the columns so that the columns after the first one have close to equimolal feed rates thus reducing the costs of those columns.

Feed Composition. When the feed data are given in weight rate and weight fractions, the mass balances on a column can be carried out on weight basis. However it is important to convert all compositions and rates to molar basis before proceeding with the design calculations. Column simulators allow the specification of the feed composition in a variety of bases—mole fractions, weight fractions, volume fractions, and molar or weight component flows. The engineer should use the basis on which the data are available to avoid unnecessary conversion calculations that may introduce errors into the simulation.

Feed Enthalpy. The feed enthalpy has a major effect on how the liquid and vapor rates in the column change at the *feed tray*, the tray where the feed is introduced into the column. It also affects the optimum location of the feed tray in the column, as we shall see later. The important characteristic of the feed enthalpy is its value relative to its saturated liquid—bubble point—and vapor enthalpies at the pressure on the feed tray.

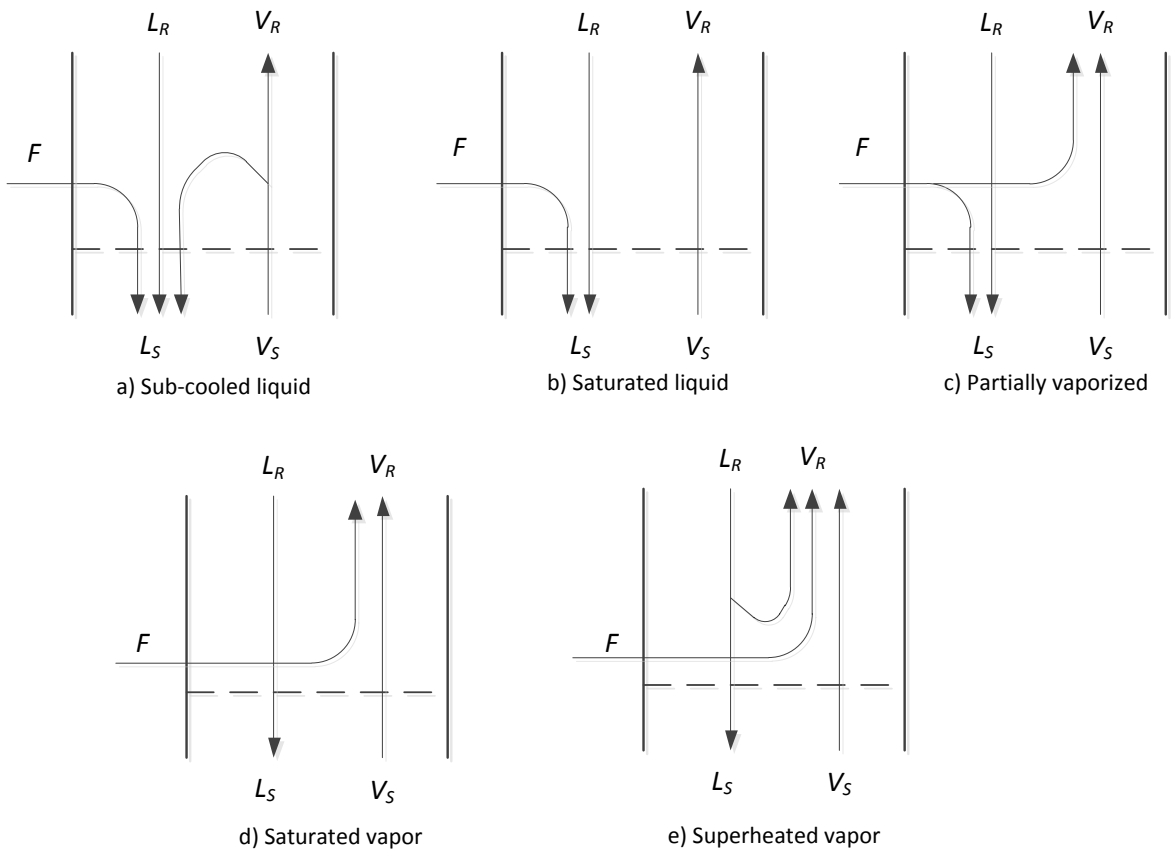


Figure 1. Effect of feed enthalpy on liquid and vapor rates

Figure 1 shows what happens at the feed tray for different feed enthalpy conditions.

a) **Sub-cooled liquid.** If the feed is below its bubble point when it enters the feed tray (at the tray pressure), some of the vapors coming into the feed tray must condense at the feed tray to bring the feed up to boiling. So there is a reduction in the vapor rate at the feed tray and the liquid rate increases by the flow of the feed plus the rate of vapors condensed on the tray.

b) **Saturated liquid.** If the feed is at its bubble point as it enters the feed tray, the liquid rate increases by the flow of the feed and that vapor rate does not appreciably change as it goes through the feed tray.

c) **Partially vaporized.** If the feed is partially vaporized the liquid rate increases by the fraction of the feed that is liquid and the vapor rate increases by the fraction of the feed that is vapor.

d) **Saturated vapor.** If the feed is at its dew point as it enters the feed tray the vapor rate increases by the feed rate and the liquid rate does not appreciably change.

e) **Superheated vapor.** If the feed is above its dew point as it enters the feed tray some on the liquid entering the tray must evaporate to cool the feed down to saturation. The liquid rate is reduced by the fraction that vaporizes and the vapor rate increases by the feed rate plus the rate of vaporization of the liquid.

So, the feed enthalpy has an important effect on the design of the column, as we shall see in more detail later. For now let us point out that a sub-cooled feed requires the generation of a higher vapor rate in the reboiler of the column, and a superheated feed requires the condensation of additional vapor in the condenser. Both of these extremes are usually avoided by either pre-heating or pre-cooling the feed.

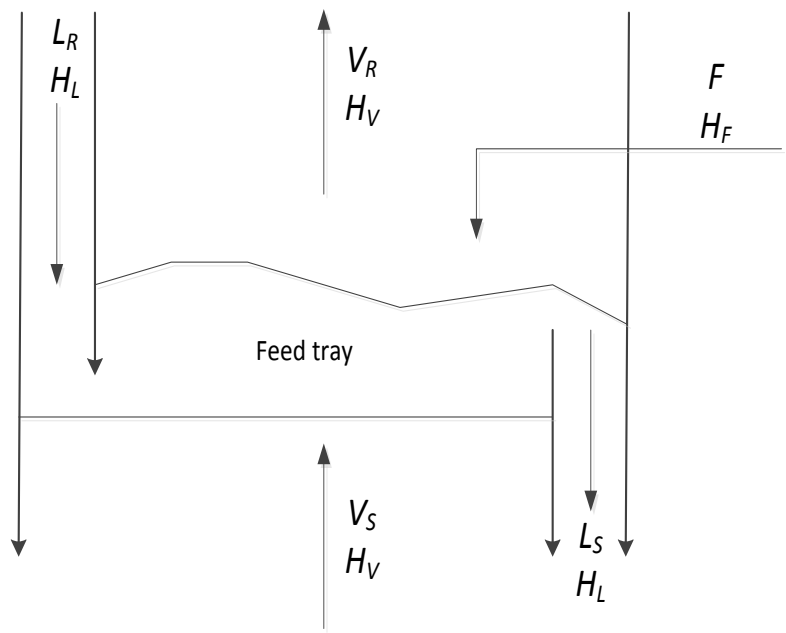


Figure 2. Enthalpy balances on feed tray

A clever method of representing the five enthalpy conditions just described is the definition of the *feed enthalpy condition* parameter *q*. It is essentially the fraction of the feed that is liquid at the pressure of the feed tray. To derive it we do an enthalpy balance on the feed tray, Fig. 2. Let us make the simplifying assumptions that, as the composition and temperature do not change much around the feed tray, the specific enthalpies of the liquid and vapor streams do not change much as they flow through the feed tray, and as the tray is assumed to be an equilibrium stage at a composition near that of the feed, they are equal to the liquid and vapor saturation enthalpies of the feed. Let us further assume equimolal overflow so that the liquid and vapor rates do not change from tray to tray in the sections above and below the feed tray; they change only at the feed tray. An enthalpy balance on the tray of Fig. 1 gives:

$$L_R H_L + V_S H_V + F H_F = L_S H_L + V_R H_V \qquad (1)$$

1 eqn., 2 unks. [L_S, V_S]

Where L and V are the liquid and vapor rates (kgmole/hr), respectively, and subscripts R and S refer to the column sections above and below the feed tray. H_L and H_V are the liquid and vapor molar enthalpies (J/kgmole), respectively, assumed not to change across the feed tray and to be near the saturation enthalpies of the feed. F is the feed rate and H_F its enthalpy. A total mole balance on the tray gives

$$L_R + V_S + F = L_S + V_R \tag{2}$$

2 eqns., 2 unks. Solved!

Eliminate V_S from the mole balance, substitute into the enthalpy balance, and solve for L_S:

$$L_S = L_R + \frac{H_V - H_F}{H_V - H_L} F \tag{3}$$

Define the *feed enthalpy condition* parameter:

$$q = \frac{H_V - H_F}{H_V - H_L} \tag{4}$$

Using this parameter, the change in the liquid and vapor rates can be calculated:

$$\begin{aligned} L_S &= L_R + qF \\ V_S &= V_R - (1-q)F \end{aligned} \tag{5}$$

The five different enthalpy conditions of the feed can now be represented in terms of parameter q:

(a) Sub-cooled liquid: $H_F < H_L$, $q > 1$

(b) Saturated liquid: $H_F = H_L$, $q = 1$

(c) Partially vaporized: $H_V > H_F > H_L$, $0 < q < 1$

(d) Saturated vapor: $H_F = H_V$, $q = 0$

(e) Superheated vapor: $H_F > H_V$, $q < 0$

Parameter q is very useful for representing the enthalpy condition of the feed. Later we will use it in the graphical solution of the column design equations.

It is important that the feed be fed to the column at a point where the composition on the feed tray is close to the composition of the feed to avoid undoing the work of separation by mixing streams of different compositions. This is why there is an optimum feed tray location. The enthalpy condition of the feed determines whether the composition to be matched on the feed tray is that of the liquid, the vapor, or somewhere between the two.

Why must we use molar units to do the distillation calculations?

There are two reasons for using molar rates and mole fractions in distillation calculations instead of weight rates and weight fractions. The first one is that the equilibrium relationships such as Raoult's Law apply only when using mole fractions. The other one is that the vapor and liquid rates vary less from tray to tray when expressed in molar units (kgmole/hr). The latent heats of vaporization of most substances on a molar basis, J/kgmole, are functions of their normal boiling points in absolute units (K). Because the normal boiling points of the components in a distillation column are usually near each other, their latent heats of vaporization are also close to each other and, as the energy effects on the trays are mostly due to vaporization and condensation—sensible heat effects are very small—the vapor rates on molar basis do not vary much from tray to tray. From the total molar balance, Eq. 3.1, we see that if the vapor rates don't vary, the liquid rates do not vary much

either. This lets us simplify the solution of the equations by assuming *equimolal overflow*: assume the vapor and liquid rates remain the same from tray to tray, except for the feed tray, the tray where the feed is introduced into the column. By assuming equimolal overflow we do not have to do an enthalpy balance on each tray or calculate the temperature or do flash calculations on each tray, provided we have independent access to the equilibrium data.

2. Column Specifications

Figure 3 shows a sketch of a distillation column with two product streams, the *distillate* and the *bottoms product*. The feed enters at the *feed tray* and the section of the column above the feed tray is called the *rectifying* section. Its purpose is to purify the vapors flowing up the column by contacting them in the trays with the liquid coming down the column to remove the high-boiling components from the vapor. The section consisting of the feed tray and the trays below it is called the *stripping* section. Its purpose is to strip off the low-boiling components from the liquid flowing down the column by contacting them in the trays with the vapor flowing up the column.

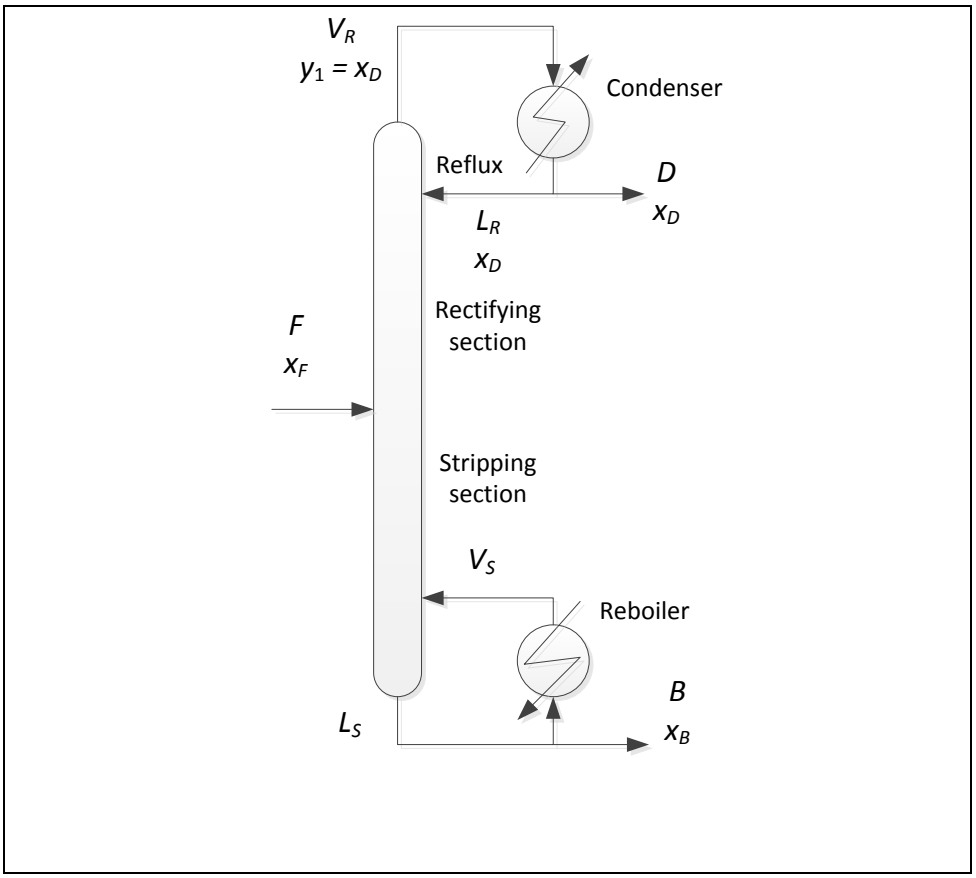


Figure 3. Schematic of a typical distillation column

The vapors coming out of the top of the column are either partially or totally condensed in the *condenser* and part of the resulting liquid is returned to the column to provide the liquid for the rectifying section. The returned liquid is called the *reflux*. The liquid coming out of the bottom of the column is partially vaporized in the *reboiler* to provide the vapor for the stripping section.

The five degrees of freedom or design variables on a distillation column are the two separation specifications, usually the two product purities, the ratio of the reflux rate to the distillate product rate, known as the *reflux ratio*, the column operating pressure, and the specification that the feed must be fed at the optimum location. Let us look at each in turn.

2.1 Separation Specifications.

The two separation specifications determine how pure the products are and how much of each component is recovered. In industry the purity of the products is usually expressed as the maximum allowed impurity or fraction of the other component. For example, instead of specifying that the distillate have 99.9% of the low-boiling or *light* component, it is specified that it contain no more than 0.1% of the high-boiling or *heavy* component. This is good practice and should always be followed.

2.2 Column Molar Balances.

Together with the feed flow and composition, the two separation specifications determine the flows of the distillate and bottoms products from the total and component molar balances; these balances are, for the binary distillation column with two products of Fig. 3:

$$\begin{aligned} F &= D + B \\ Fx_F &= Dx_D + Bx_B \end{aligned} \tag{6}$$

2 eqns., 4 unks. [D, B, x_D, x_B]

Where D and B are the molar rates of the distillate and bottoms products (kgmole/hr), respectively, x is the mole fraction, usually of the low-boiling component, and subscripts D and B refer to the distillate and bottoms product, respectively.

The solution of Eqs. 6 requires two separation specifications. These specifications can be the compositions of the distillate and bottoms products, x_D and x_B which case the equations to be rearranged to calculate the product rates:

$$D = \frac{x_F - x_B}{x_D - x_B} F \qquad (7)$$
$$B = F - D$$

Another way to specify the separation is the *recovery* of the most important component and its composition in the product stream. The recovery is the fraction of the component flowing in with feed that is recovered in the product stream. The recovery of the low-boiling component is Dx_D/Fx_F, and that of the high-boiling component is $B(1 - x_B)/F(1 - x_F)$. Here we have assumed that the mole fractions are those of the low-boiling component. The following example illustrates the options on specifying the separation in the column.

Example 1. Column Material Balances

A distillation column is fed 750 kgmole/hr of a mixture containing 35 mole% benzene and the balance toluene.

(a) Determine the distillate and bottoms product rates if the distillate must contain no more than 0.1 mole% toluene and the bottoms must contain no more than 0.5 mole% benzene.

From Eq. (7), with $x_D = 1 - 0.001 = 0.999$:

$$D = \frac{0.35 - 0.005}{0.999 - 0.005} 750 \frac{\text{kgmole}}{\text{hr}} = 260 \frac{\text{kgmole}}{\text{hr}}$$
$$B = 750 - 260 = 490 \frac{\text{kgmole}}{\text{hr}}$$

The recovery of benzene is

$$(0.999)(260)/(0.35)(750) = 0.989$$

(b) Determine the distillate and bottoms product rates to recover 98% of the benzene in the distillate containing no more than 0.1 mole% toluene.

The rate of benzene into the column is (0.35)(750) = 262 kgmole/hr. From the recovery, the rate of benzene recovered in the distillate is (0.98)(262) = 255 kgmole/hr. The product rates, with x_D = 1 – 0.001 = 0.999, are:

$$D = \frac{255}{0.999} = 255\,\frac{\text{kgmole}}{\text{hr}} \qquad B = 750 - 255 = 495\,\frac{\text{kgmole}}{\text{hr}}$$

(c) Determine the distillate and bottoms product rates to recover 99% of the toluene in the bottoms product containing no more than 0.5 mole% benzene.

The rate of toluene in the feed is (1 – 0.35)(750) = 488 kgmole/hr. From the recovery, the rate of toluene recovered in the bottoms product is (0.99)(488) = 483 kgmole/hr. The flow of the products is then:

$$B = \frac{483}{1 - 0.005} = 485\,\frac{\text{kgmole}}{\text{hr}} \qquad D = 750 - 485 = 265\,\frac{\text{kgmole}}{\text{hr}}$$

(d) If the distillate rate is set at 300 kgmole/hr, what is the maximum possible purity of the distillate?

The rate of benzene into the column is (0.35)(750) = 262 kgmole/hr. The maximum composition of the distillate with a fixed rate occurs when all of the benzene is recovered in the distillate. The composition is then

$$x_{D\max} = \frac{262}{300} = 0.875$$

this is a very low purity for the product.

(e) If the distillate rate is set at 200 kgmole/hr, what is the maximum possible recovery of the benzene?

The maximum possible benzene recovery with this fixed distillate flow occurs when the distillate is 100 mole% benzene. The recovery is then

$$\text{Benzene recovery} = \frac{200}{262}100\% = 76.3\%$$

this is a very low recovery.

The last two parts of the preceding example show that it is not a good idea to specify the product rates from a distillation column because either a low product purity results when the specified rate is too high, or a low recovery results when the specified rate is too low. The reason is that, by the conservation of component mass, the column cannot produce more or less of the component that goes into it. Notice that, by the conservation of total moles, the same thing would happen if the rate of the bottoms product were to be specified. This principle applies when simulating the column with a process simulator, and when operating the column after it is built. That is, it is not a good idea to arbitrarily control the flow of the products from the column.

2.3 Reflux Ratio.

In addition to the two separation specifications, the engineer must specify the ratio of the flow of liquid returned to the column from the condenser to the flow of the distillate product. This *reflux ratio* has a significant effect on the number of equilibrium stages required to do a given separation and on the vapor flow rate in the column, therefore the size and cost of the column significantly depends on the specified reflux ratio. The reflux ratio is defined by:

$$R_D = \frac{L_R}{D} \tag{8}$$

Where R_D is the reflux ratio, L_R is the reflux rate (kgmole/hr), and D is the distillate rate (kgmole/hr). A balance around the condenser (see Fig. 3) gives us the rate of vapors leaving the column, V_R (kgmole/hr):

$$V_R = L_R + D = (R_D + 1)D \tag{9}$$

Recalling that the distillate rate D depends on the feed rate and composition and the separation specifications, we see from Eqns. (8) and (9) that the liquid and vapor rates in the column increase with the reflux ratio. The higher rates require a larger column diameter and a higher heat duty for the condenser. As the molar rates of liquid and vapor vary little from tray to tray, the liquid and vapor rates in the rectifying section of the column are near L_R and V_R, respectively. From Eq. (5), the liquid and vapor rates in the stripping section of the column, L_S and V_S, depend on the corresponding rates in the rectifying section. Therefore, all of the liquid and vapor rates in the column, and consequently the reboiler heat duty, increase when the reflux ratio is increased.

As we shall soon see, increasing the reflux ratio reduces the number of equilibrium stages required for a given separation. This is because the ratio of liquid to vapor rate in the rectifying section increases with the reflux ratio allowing the removal of more of the high-boiling components from the vapor at each tray.

2.4 Boil-up Ratio.

Sometimes the *boil-up ratio* is used instead of the reflux ratio specification to design the column. The boil-up ratio is defined by

$$R_B = \frac{V_S}{B} \tag{10}$$

Where R_B is the boil-up ratio, V_S is the rate of vapors entering the column (kgmole/hr), and B is the bottoms product rate (kgmole/hr). A balance around the reboiler (see Fig. 3) gives

$$L_S = V_S + B \tag{11}$$

Where L_S is the liquid rate leaving the bottom of the column (kgmole/hr).

Notice that since the liquid and vapor rates in the stripping section are related to the liquid and vapor rates in the rectifying section by the balances on the feed tray, Eq. (1), the boil-up ratio is not independent of the reflux ratio and both cannot be specified on the same column.

When the reflux ratio is increased the boil-up ratio also increases, because the vapor rate in the stripping section increases. We see from Eq. (11) that the vapor rate is always

greater than the liquid rate in the stripping section, so when the reflux and boil ratios increase, the ratio of vapor to liquid rate in the stripping section increases. This strips more of the low-boiling components from the liquid flowing down the column at each tray and reduces the number of trays required to do a given separation. However, the higher vapor rate increases the required column diameter and the heat duty of the reboiler and consequently the size and cost of the column.

2.5 Pressure Specification.

The pressure in the column determines the equilibrium relationships in the column, that is, the relationship between the vapor and liquid compositions at equilibrium on each tray. Most process simulators perform the calculations assuming that each tray is an *ideal* or *equilibrium* stage. In such a stage the vapor and liquid streams leave in equilibrium with each other. The change in the composition of the vapor as it flows through an equilibrium stage is,

$$y_{i,k} - y_{i,k+1} = K_i(T_k, P, x_{i,k})x_{i,k} - y_{i,k+1} \qquad (3.17)$$

where y and x are the vapor and liquid mole fractions, respectively, K is the equilibrium K-factor, T is the temperature, P is the pressure, subscript i refers to component i and subscript k refers to the tray number. We see that the equilibrium K-factor is a function of the column operating pressure P.

For most mixtures the lower the pressure the higher the relative volatility—ratio of the K-factors of the components—and therefore the easier the separation, that is, fewer required trays and/or smaller reflux ratios. This is one reason the

pressure must not be specified higher than it needs to be. However, the most important consideration in selecting the column pressure is its effect on the column temperatures. As the liquid on each tray is at or near equilibrium with the vapor, its temperature is near its bubble point which is the boiling point of the mixture on the tray. This applies to the column condenser which, assuming no sub-cooling of the condensate, is near the bubble point of the distillate, and to the reboiler which is at the bubble point of the bottoms product. Therefore, as the pressure is increased, the temperatures on all the trays, the condenser and the reboiler increase. So, there are two important considerations when selecting the column operating pressure:

- The pressure must be selected so that the condenser temperature is high enough to avoid unnecessary use of refrigeration as the cooling medium in the condenser, as refrigeration is a much more expensive coolant than cooling tower water or air. As we shall see later the minimum temperature of operation of the condenser is about 45°C (110°F).
- The pressure must not be so high that it requires a higher pressure steam than necessary to produce the vapors in the reboiler, because higher pressure steam is more costly.

Which of these considerations restricts the pressure in the column depends on the normal boiling points of the components being separated and the purities of the distillate and bottoms products. When the normal boiling point of the low-boiling component is around ambient temperature, the column may need to be operated under pressure to insure the condenser temperature is above 45°C. If the normal boiling points of the components is

much higher than ambient temperature the reboiler temperature becomes the limiting consideration and the column may have to be operated under vacuum to avoid the used of unnecessarily expensive high-pressure steam or fuel in a heated reboiler. If the normal boiling point of the low-boiling component is far below ambient temperature, it may not be possible to avoid using refrigeration.

Another important consideration is that the reboiler must operate at a higher pressure than the condenser to overcome the pressure drop experienced by the vapors as they flow through the trays up the column. Normally a pressure drop of about 1.5 kPa per ideal tray is allowed in the design. Notice that this means the pressures in the condenser and the reboiler are not independent of each other, as the difference between them depends on the number of trays required for the separation.

2.6 Optimum Feed Tray Location.

The fifth and final specification in the design of a distillation column is that the feed be introduced into the column at its optimum location. As the purpose of the distillation column is to separate the components in the feed, any mixing of streams of different composition causes the undoing of the work of separation. The optimum feed tray location is the one where the feed is of about the same composition as the contents of the tray. Notice that this may depend on the enthalpy condition of the feed. For example, if the feed is saturated liquid at the pressure in the feed tray it must be at the same composition as the liquid on the tray, but if it is a saturated vapor feed, it must be at the same composition as the vapor leaving the feed tray. At other feed enthalpy conditions the composition of the feed must be near the

average composition of the liquid and vapor on the feed tray, weighted by the feed enthalpy condition parameter q.
As we shall soon see, introducing the feed at the wrong tray requires more trays than necessary to do the separation.

3. Number of Required Equilibrium Stages by McCabe-Thiele

In 1925 two chemical engineering graduate students, McCabe and Thiele (McCabe & Thiele, 1925), developed a graphical method to determine the number of equilibrium stages required in a binary distillation column. Although today we have powerful simulation programs to carry out the detailed distillation calculations, the McCabe-Thiele method still provides insight into the effect of the various design specifications on the size of a column. This is why we are presenting it here.
The McCabe-Thiele method uses the x-y diagram shown in Fig. 4. The diagram shows the equilibrium line as the mole fraction y of the more volatile component in the vapor versus the mole fraction x of the same component in the liquid. Because it is the more volatile or low-boiling component the equilibrium line falls above the diagonal or $y = x$ line, also shown in the diagram. Recall that the position of the equilibrium line depends on the column operating pressure; the lower the pressure the farther the equilibrium line from the diagonal line. To determine the required number of equilibrium stages, we must draw the two column *operating lines* to the diagram, one for the rectifying section and another for the stripping section.

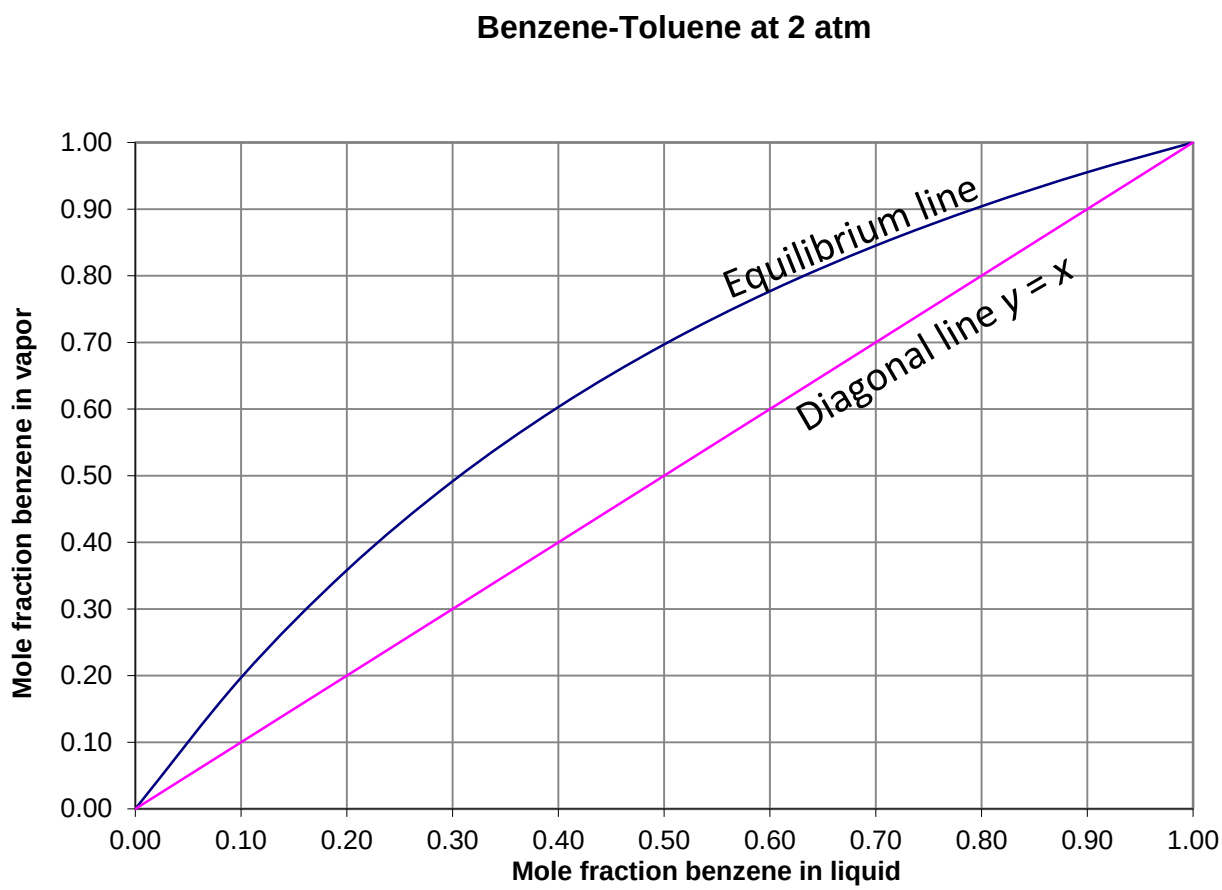


Figure 4. *x-y* diagram for benzene-toluene

3.1 Rectifying Operating Line.

The operating lines are obtained from material balances around sections of the column. For the rectifying operating lines we write total and molar balances around the top *k* trays of the column, above the feed tray, as Fig. 5 shows:

$$\begin{aligned} V_{k+1} &= L_k + D \\ V_{k+1} y_{k+1} &= L_k x_k + D x_D \end{aligned} \qquad (13)$$

Where V, L, and D are the molar rates of vapor, liquid, and distillate (kgmole/hr), respectively, x and y are the mole fractions of the low-boiling component in the liquid and vapor, respectively, and subscript k refers to the kth plate in the rectifying section.

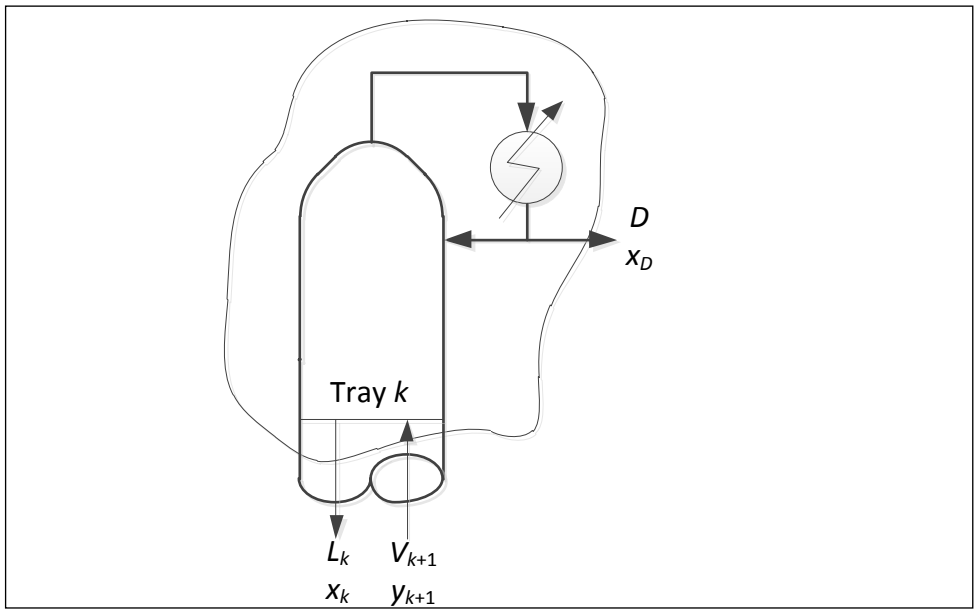


Figure 5.

Balances around rectifying section

Solving for y_{k+1} gives us the equation of the rectifying operating line,

$$y_{k+1} = \frac{L_k}{V_{k+1}} x_k + \frac{D x_D}{V_{k+1}} \tag{14}$$

If we now assume equimolal overflow, the liquid and vapor rates do not change throughout the rectifying section so that the operating line becomes a straight line; let $L_k = L_R$, $V_{k+1} = V_R$, and substitute:

$$y_{k+1} = \frac{L_R}{V_R} x_k + \frac{D x_D}{V_R} \tag{15}$$

Notice that the operating line relates the compositions of the streams that cross each other, that is, the vapor entering a tray with the liquid exiting the same tray. On the other hand the equilibrium line relates the composition of the streams leaving an equilibrium stage.

It is convenient to express the operating line equation in terms of the reflux ratio. To do this we substitute Eqs. (8) and (9) into Eq. (15) and simplify:

$$y_{k+1} = \frac{R_D}{R_D+1} x_k + \frac{x_D}{R_D+1} \tag{16}$$

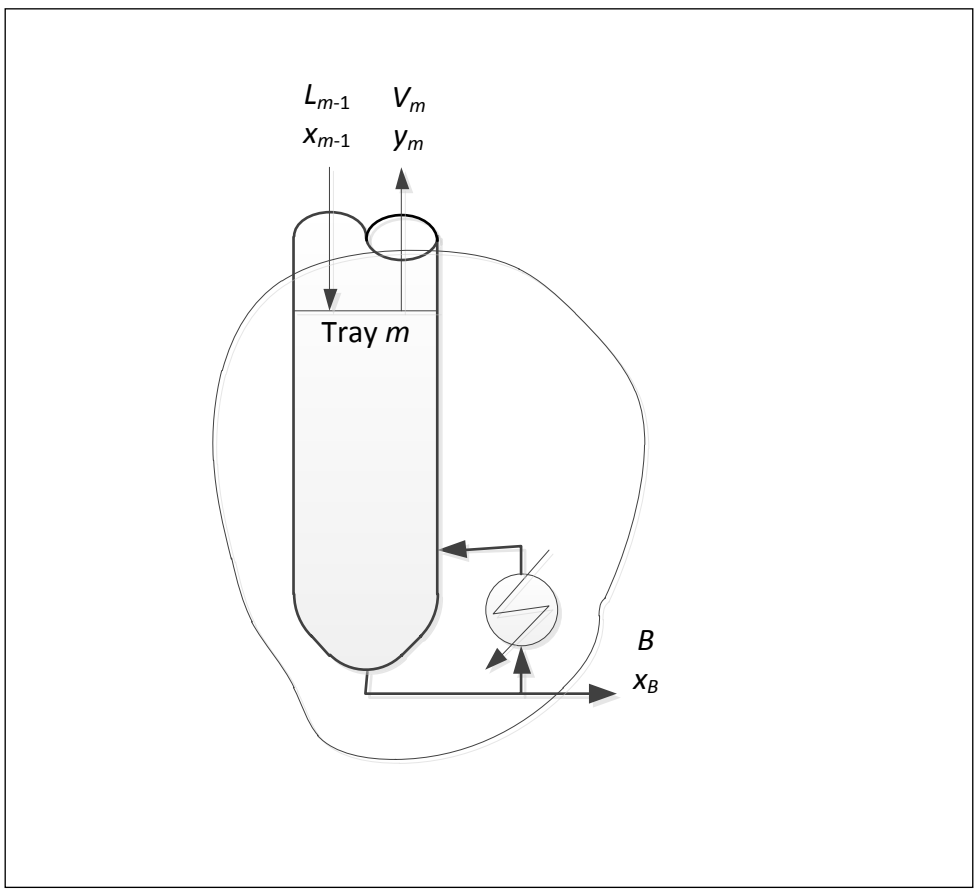


Figure 6. Balances around stripping section

3.2 Stripping Operating Line.

To obtain the stripping operating line we write molar and component balances around the bottom of the column, below the feed tray, as sketched in Fig. 3.7:

$$V_m = L_{m-1} - B$$
$$V_m y_m = L_{m-1} x_{m-1} - B x_B \qquad (17)$$

Where L, V and B are the molar rates of the liquid, vapor and bottoms product (kgmole/hr), respectively, y and x are the mole fractions of the low-boiling component, and subscript m refers to the mth tray in the stripping section of the column. Solving for y_m gives us the equation of the stripping operating line:

$$y_m = \frac{L_{m-1}}{V_m} x_{m-1} - \frac{B x_B}{V_m} \qquad (18)$$

Assuming equimolal overflow, the liquid and vapor rates do not change from tray to tray and we can write:

$$y_m = \frac{L_S}{V_S} x_{m-1} - \frac{B x_B}{V_S} \qquad (19)$$

Where L_S and V_S are the liquid and vapor rates (kgmole/hr), respectively, in the stripping section of the column. So assuming equimolal overflow, both operating lines are straight.

3.3 Plotting the Operating Lines.

As the operating lines are straight, only two points are required to plot each line. For the rectifying line, from Eq. (17), these two points are:

At $x = x_D$ $\quad y = x_D$

At $x = 0$ $\quad y = \frac{x_D}{R_D+1}$

The first point says that the line crosses the diagonal line at the distillate mole fraction and the second one is the intercept of the line on the y-axis.

One point on the stripping operating line, given that $V_S = L_S - B$, is

$$At\ x = x_B \qquad y = x_B$$

So the stripping line crosses the diagonal line at the mole fraction of the bottoms product. As the stripping liquid and vapor rates are related to the rectifying rates through the feed tray balances, the other point on the stripping line depends on the composition and enthalpy condition of the feed. To obtain it we must determine the *feed line*.

Feed Line.

The feed line is the location of the points where the rectifying and stripping lines intercept each other, that is, the points on the feed line lie on both operating lines. Rearranging Eqs. (15) and (16):

$$\begin{aligned} V_R y &= L_R x + D x_D \\ V_S y &= L_S x - B x_B \end{aligned} \tag{20}$$

Subtract the second equation from the first one to obtain:

$$(V_R - V_S)y = (L_R - L_S)x + D x_D + B x_B \tag{21}$$

Substitute Eqs. (5) and (7) to obtain:

$$(1-q)Fy = -qFx + Fx_F \tag{22}$$

Simplify and solve for y to obtain the equation of the feed line:

$$y = -\frac{q}{1-q}x + \frac{x_F}{1-q} \tag{23}$$

From the equation, at $x = x_F$, $y = x_F$, so the feed line crosses the diagonal line at the mole fraction of the feed with a slope of $-q/(1 - q)$. The slope depends on the enthalpy condition of the feed, as follows:

(a) Sub-cooled liquid feed, $q > 1$, slope > 1
(b) Saturated liquid feed, $q = 1$, line is vertical
(c) Partially vaporized feed, $0 < q < 1$, slope < 0
(d) Saturated vapor feed, $q = 0$, line is horizontal
(e) Superheated vapor feed, $q < 0$, 0 < slope < 1

Sample feed lines are shown for each of these conditions on Fig. 7. When the line is not vertical or horizontal it is plotted by calculating a convenient point using Eq. (23). Depending on the value of q a convenient point is obtained by calculating x at either $y = 0$ or $y = 1$, or the intercept of the y-axis (y at $x = 0$).

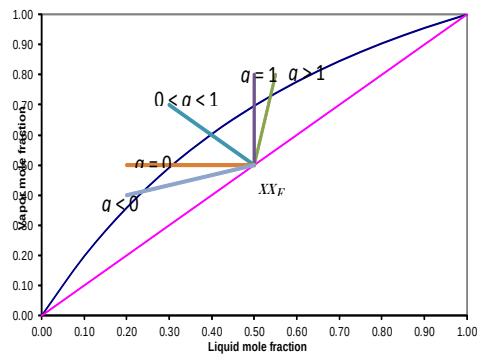


Figure 7. Samples of feed lines on *x-y* diagram

We can now complete the plotting of the operating lines. The procedure is as follows:

1. Mark the three points on the diagonal at x_B, x_F and x_D.
2. Plot the rectifying operating line between the point on the diagonal at x_D and the y-axis intercept, $x_D/(R_D + 1)$.
3. Plot the feed line.
4. Plot the stripping line between the point on the diagonal at x_B and the point where the rectifying operating line crosses the feed line.

These lines are shown on the x-y diagram of Fig. 8, with all the points used to plot the lines marked. On the plot the feed line

is assumed to be partially vaporized. Once the operating lines are plotted, the number of equilibrium stages can be determined.

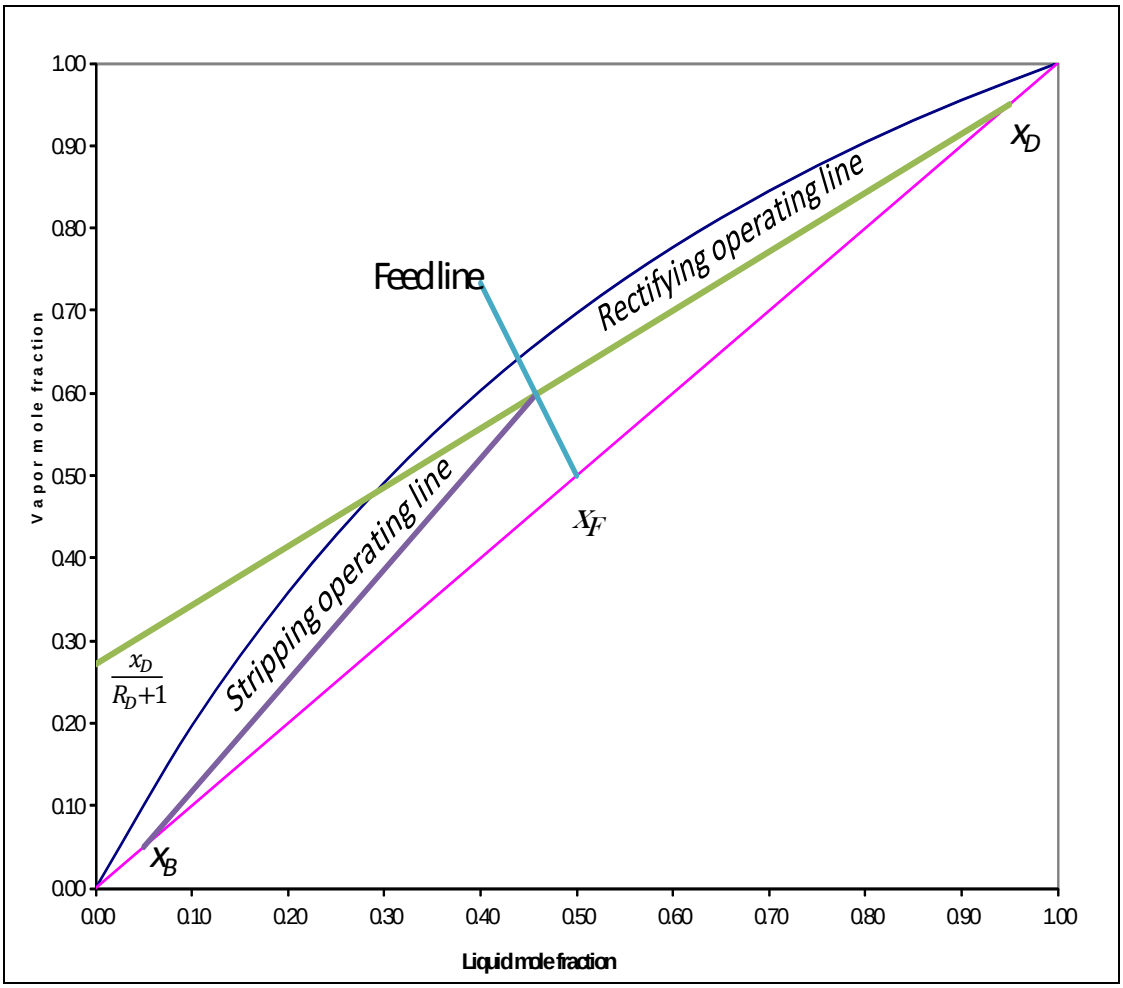


Figure 8. Drawing operating lines on *x-y diagram*

3.4 Determining the Required Number of Equilibrium Stages.

Let us review the steps that have brought us to the plot of Fig. 8:

- With the feed composition x_F and the two separation specifications we determined x_D and x_B.
- At the selected column operating pressure we have obtained the data to plot the equilibrium line.
- We have selected an appropriate reflux ratio.

- From the enthalpy of the feed we have determined the feed enthalpy parameter q.

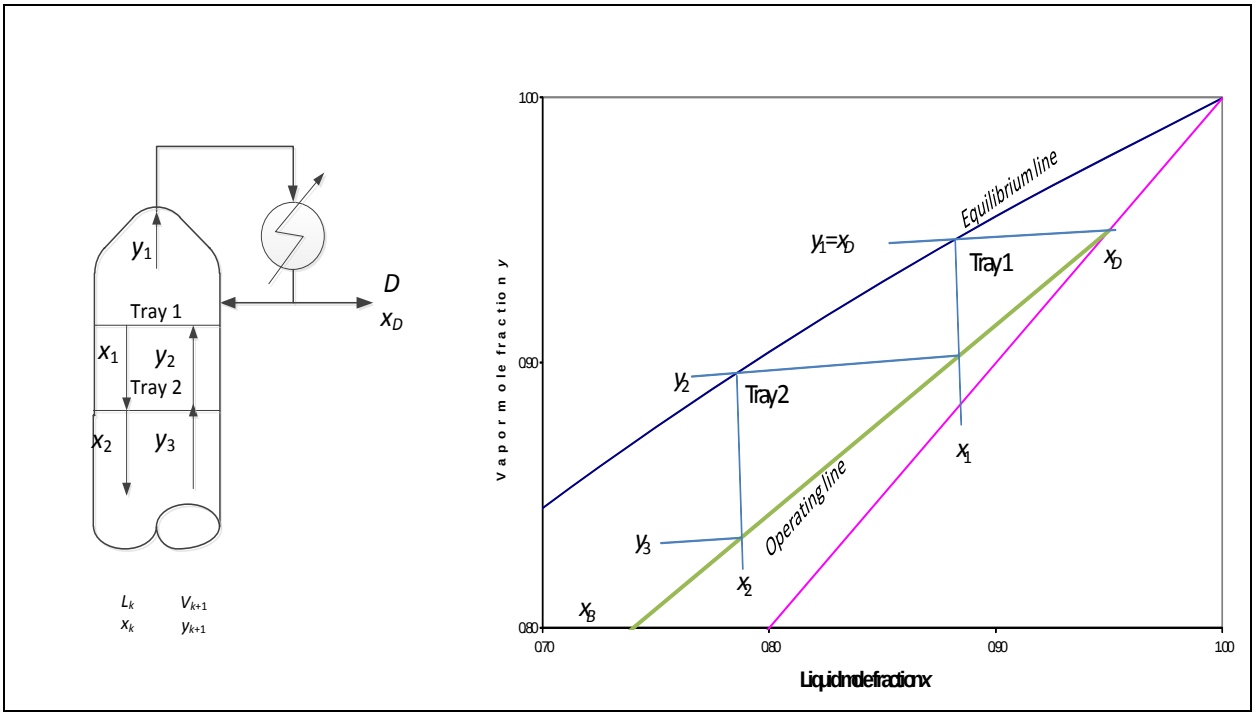


Figure 9. Stepping the compositions from tray to tray

When all the vapors leaving the top of the column are condensed, the vapor leaves the top of the column at the same composition as the distillate.

1. As the vapor leaves the top tray, tray 1, its composition is y_1 (see Fig. 9):

$$y_1 = x_D$$

2. The liquid leaving the first tray is in equilibrium with the vapor leaving the tray, as it is assumed to be an ideal tray, so the point (x_1, y_1) is on the equilibrium line. To obtain this point we draw a horizontal line at $y = x_D$ from the diagonal line to the equilibrium line, as in Fig. 9.
3. The rectifying operating line relates the composition of the vapors entering tray 1, y_2, to the composition leaving.

So the point (x_1, y_2) is on the operating line. To obtain it we draw a vertical line at $x = x_1$ from the equilibrium line to the operating line, as in Fig. 9.

4. We continue this process to obtain point (x_2, y_2) on the equilibrium line by drawing a horizontal line from point (x_1, y_2), then a vertical line to obtain point (x_2, y_3) on the operating line and so on, as we move down the column.

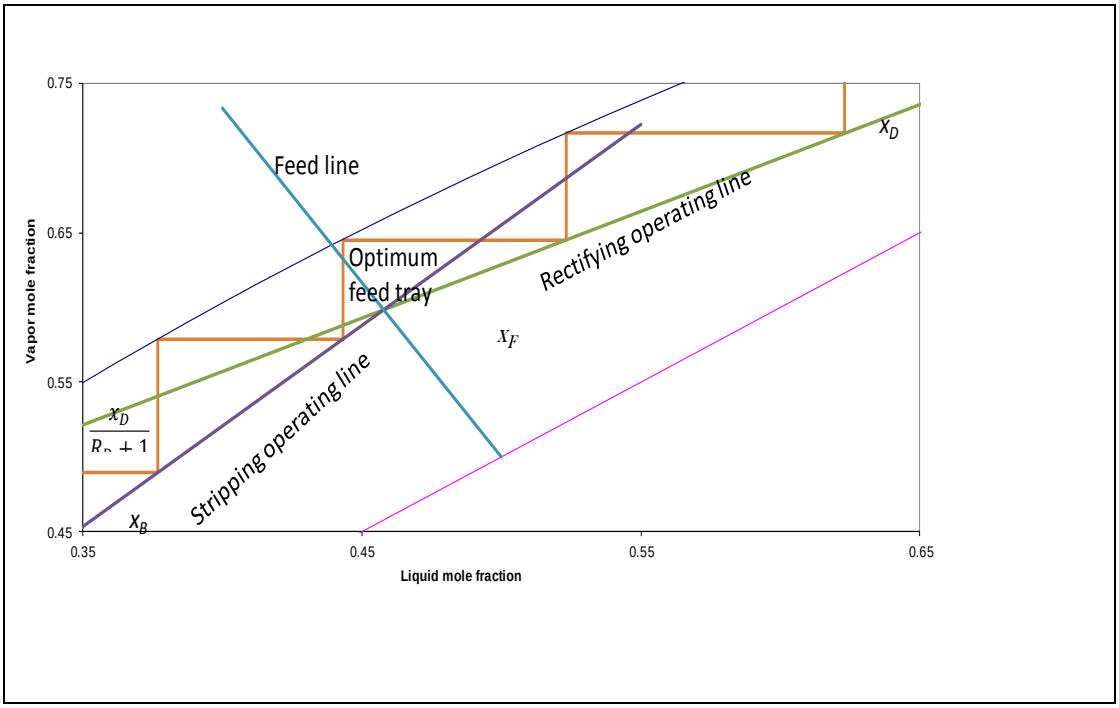


Figure 10. At optimum feed tray switch operating lines

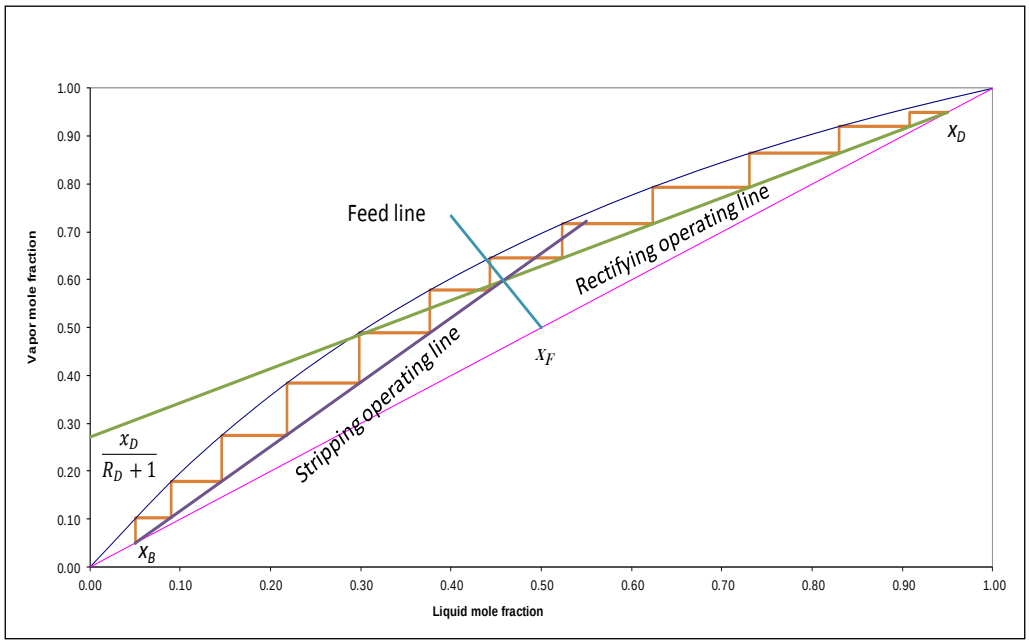
5. When we near the feed line we have the option of moving down to either the rectifying or the stripping operating line (see Fig. 10). It is easy to see that on the right side of the feed line there is a higher change in the vapor composition when the rectifying line is selected instead of the stripping line. However, to the left of the feed line a higher change in vapor composition occurs when the stripping operating line is selected. Selecting the stripping operating line means that the feed has been introduced on the tray where the switch is made, that is,

the optimum feed tray location is the tray where the two operating lines and the feed line cross.

6. We continue stepping down the column until $x_N < x_B$.

The completed McCabe-Thiele procedure is shown in Fig. 11. The plotting of the stages can be done either from the top down starting at the top tray, as described above, or from the bottom up starting at the reboiler. In the figure the feed has been assumed to be partially vaporized.

Figure 11. Complete McCabe-Thiele diagram



Example 2. Benzene-Toluene Column.

A column is fed 550 kgmole/hr of a saturated liquid containing 45 mole% benzene and the balance toluene. It is desired to recover 95% of the benzene in a distillate product containing no more than 2 mole% toluene. Determine the vapor and liquid rates and the required number of equilibrium stages with a reflux ratio of 1.95. The column operates at 2 atm pressure (203 kPa). Equilibrium data are available at that pressure.

Solution. The design of a binary distillation column (two components) requires eight specifications. We are given the feed flow, benzene mole fraction, and enthalpy condition; we are also given the benzene recovery, distillate composition, reflux ratio, and the operating pressure with equilibrium data. The eighth specification is implied, to have the feed enter at the optimum tray.

For 95% recovery of the benzene in the distillate with x_D =(1 − 0.02) = 0.98 mole fraction benzene:

$$D = \frac{0.95Fx_F}{x_D} = \frac{0.95(550)(0.45)}{0.98} = 240\,\frac{\text{kgmole}}{\text{hr}}$$

Total molar balance (see Fig. 3):

$$B = F - D = 550 - 240 = 310 \text{ kgmole/hr}$$

Benzene balance: $x_B = \frac{Fx_F - Dx_D}{B} = \frac{(550)(0.45)-(240)(0.98)}{310} = 0.040$

Reflux ratio: $L_R = R_D D = (1.95)(240) = 468$ kgmole/hr

Balance around condenser: $V_R = L_R + D = 468 + 240 = 708$ kgmole/hr

Balances around feed tray, saturated liquid feed (q = 1.0):

$L_S = L_R + qF = 468 + (1.0)(550) = 1{,}018$ kgmole/hr

$V_S = V_R - (1 - q)F = 708 - (0)(550) = 708$ kgmole/hr

These are all the liquid and vapor rates in the column. To determine the number of equilibrium stages required we draw the operating lines on the x-y diagram.

First mark the points at x = 0.04, x_F = 0.45, and x_D = 0.98 on the diagonal line. Then calculate the rectifying line intercept on the y-axis:

$$\frac{x_D}{R_D+1} = \frac{0.98}{1.95+1} = 0.332$$

The feed line is vertical through x_F = 0.45 because the feed is saturated liquid.

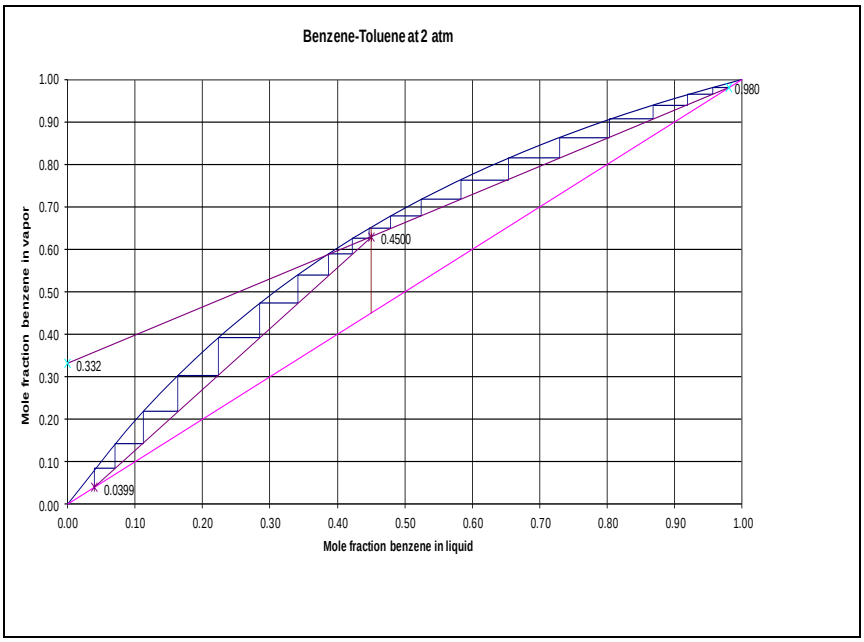


Figure 12. McCabe-Thiele diagram for Example 2

The complete McCabe-Thiele diagram, Fig. 12, shows that the separation requires 19 equilibrium stages with the feed tray on the tenth stage from the top.

In the preceding example the question was asked in terms of the required number of *equilibrium stages* instead of *ideal trays*. What's the difference? The difference arises when there is separation in the condenser and/or the reboiler. Let us look at this next.

3.5 Partial Reboiler.

Most columns have partial reboilers, that is, the liquid going into the reboiler separates into the vapor going into the column

and the bottoms liquid product. It is common to assume that the vapor leaving the reboiler is in equilibrium with the bottoms product, causing separation in the reboiler. In other words, the reboiler is an equilibrium stage and the column requires one less ideal tray.

3.6 Partial Condenser.

In a total condenser the entire vapor leaving the top of the column is condensed and the condensate is then separated into the reflux and the distillate product.

Sometimes the distillate from the columns must be removed as a vapor. The condenser is then partial because only part of the vapor is condensed and returned to the column as reflux. In such case the condenser is assumed to be an equilibrium stage and the column requires one less ideal tray.

Notice that when the condenser is total there is no separation and the vapors from the top tray have the same composition as the distillate and the reflux ($y_1 = x_D$, as assumed in a preceding discussion). With a partial condenser the reflux composition x_0 is in equilibrium with the vapor distillate (see Fig. 13), and the vapor composition y_1 is on the rectifying operating line at the composition of the liquid reflux x_0. In other words, the top stage of the column is the partial condenser.

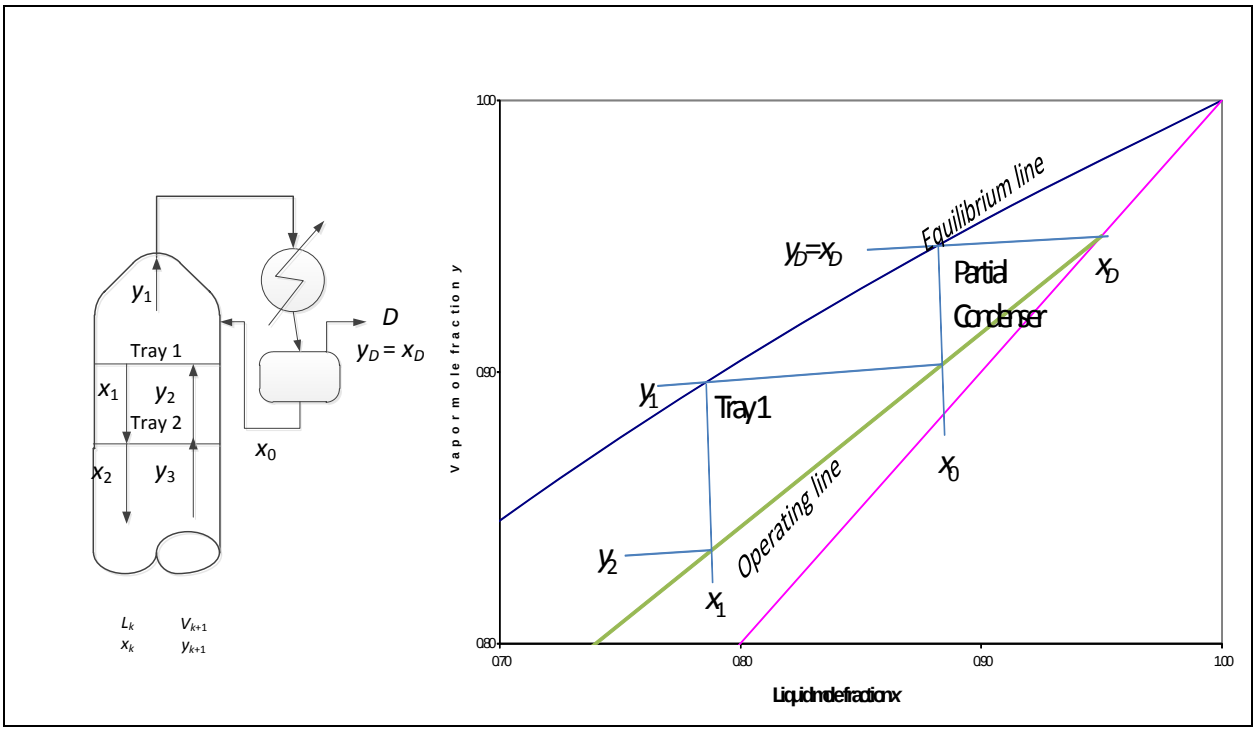


Figure 13. Partial condenser is an equilibrium stage

There are two reasons for removing the distillate as a vapor.

- One is that the distillate may be needed as a vapor in the process and it would be inefficient to condense it and then vaporize it.
- The other one is that the low-boiling components in the column may have such low normal boiling points that it would require refrigeration or very high pressures to condense them.

Some columns have two distillate products, one vapor and one liquid, that is, a vapor distillate is removed and part of the condensate is divided into the reflux and a liquid distillate. This is done when the feed to the column contains small amounts of components with very low normal boiling points (e.g., methane, hydrogen, nitrogen). These components are

removed in the vapor distillate because condensing them would require refrigeration or very high column pressures. When a column has a single vapor distillate and all the condensate is returned to the column as reflux, the condenser is said to be partial with *full reflux*.

The McCabe-Thiele procedure gives us the total number of *equilibrium stages* required for the specified separation. If the column has a partial condenser and a partial reboiler it requires two less ideal trays than equilibrium stages. If only the reboiler is partial it requires one less ideal tray than equilibrium stages.

We have seen in this section how the McCabe-Thiele graphical procedure is used to determine the number of equilibrium stages required by a distillation column. Next we will look at how the number of stages is affected by the various design specifications. Gaining this insight is the objective of this section.

4. Effect of the Design Specifications

In this section we will use the McCabe-Thiele method to help us understand how the various design specifications affect the number of equilibrium stages required for a given separation. We will look at the separation specifications, the reflux ratio, the equilibrium data, and the feed enthalpy.

4.1 Separation Specifications.

Notice in Fig. 12 that there are three areas where the operating lines are near the equilibrium line. These areas are called *pinch points* in the column because there is little change of composition from one tray to the next, so more trays are

required in these areas for a given change in composition. The middle area is at the feed tray and will be discussed with the reflux ratio. The other two areas are near the ends, the top and bottom of the column.

It is obvious that the tighter the purity specifications of the column, that is, as the specified product mole fractions approach zero at the bottom and 1.0 at the top, the closer the equilibrium line will be to the operating lines at the ends, and more stages will be required to achieve such high purities. Therefore, it is not good design practice to specify product purities or recoveries that are higher than necessary.

To determine what is a reasonable product purity specification it is necessary to look at the destination of the column products in the process. The following heuristics may serve as guidelines for selecting the separation specifications.

- If a column product, distillate or bottom, is the final process product, its purity is usually specified by the client. These purities are usually high and are sometimes in the range of parts per million of the impurity in the product.
- When the column product is recycled back to the process, the composition of the desired component in the recycle can be of the order of 1 to 5 mole% or so, because it is returned to the process and not lost. The consideration here is whether the desired component may be decomposed in the reactor, or dilute the reactants too much, or shift the equilibrium in the reactor.
- When the column product is a waste stream it is best to express the separation in terms of the recovery of the desired component in the other column product. If the waste component is a small fraction of feed to the column, a high

recovery of the desired component may result in a reasonable composition of the waste stream. For example, if the feed contains 10 mole% of the waste component and 99.5% of the desired component is recovered in the distillate, the composition of that component in the bottoms is (1 − 0.995)(0.90)/0.10 = 0.045 or 4.5 mole%, where we have assumed most of the waste component in the feed goes to the bottom when the distillate purity is high.

There is an additional consideration when there are more than two components in the feed, but we will discuss later on the multicomponent distillation document.

4.2 Reflux Ratio.

We saw in our earlier discussion that increasing the reflux ratio increases the vapor and liquid rates in the column and consequently the column cost. On the other hand, a higher reflux ratio reduces the number of stages required for a given separation and therefore the number of trays and the height of the column. There is then an optimum reflux ratio that results in the minimum total column cost. There are two limiting conditions for the reflux ratio, total reflux and minimum reflux.

Total Reflux.

At total reflux all of the vapors leaving the column at the top are condensed and returned to the column, and all of the liquid leaving at the bottom is vaporized and returned to the column. There is thus no distillates or bottoms product rates or feed to the column. Although the column cannot be designed for this condition it is sometimes operated this way during startup until the products reach near design compositions. Because $F = 0$, $D = 0$, and $B = 0$, $L_S = L_R$ and $V_S = V_R$, both operating lines become $y =$

x, that is, the operating lines coincide with the diagonal line on the x-y diagram. Figure 14 shows that the number of stages required at total reflux is the *minimum number of stages* required. Notice that this number depends only on the position of the equilibrium line and the specified product mole fractions x_D and x_B. It does not depend on the feed variables.

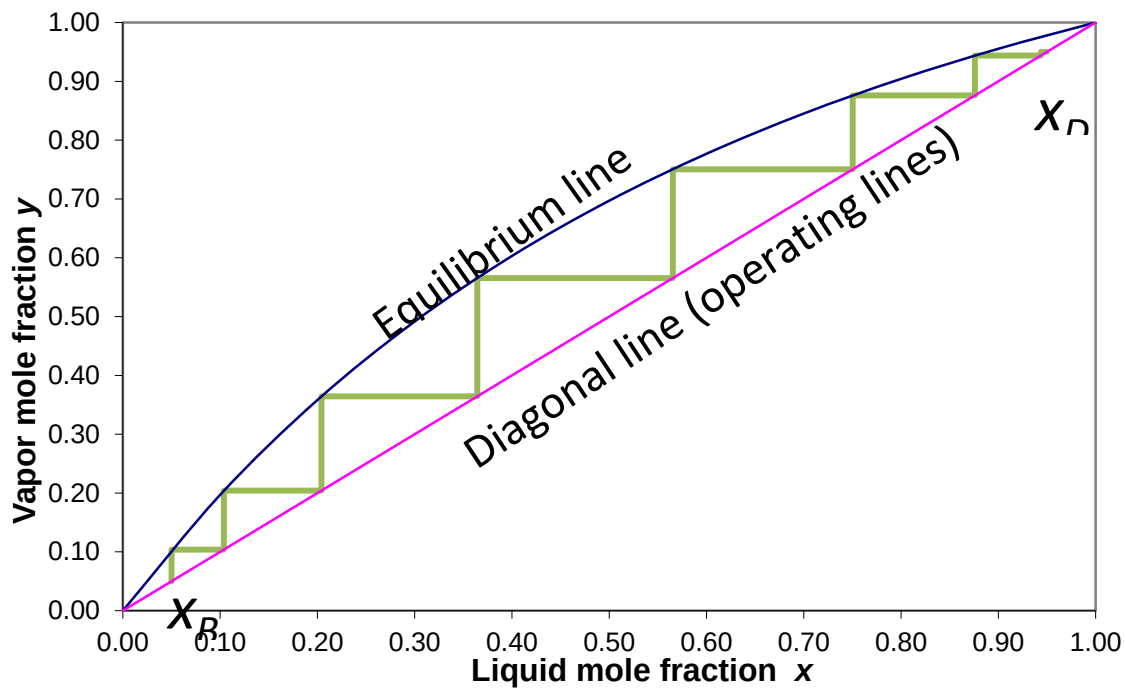


Figure 14. Minimum number of equilibrium stages at total reflux

Minimum Reflux Ratio.

As the reflux ratio is reduced the operating lines approach the equilibrium line and it takes more stages to carry out the separation. At the *minimum reflux ratio* the operating lines touch the equilibrium line. This usually happens at the point where the feed line crosses the equilibrium line as Fig. 15 shows. Notice that under this condition it would take an infinite number of stages to achieve the separation.

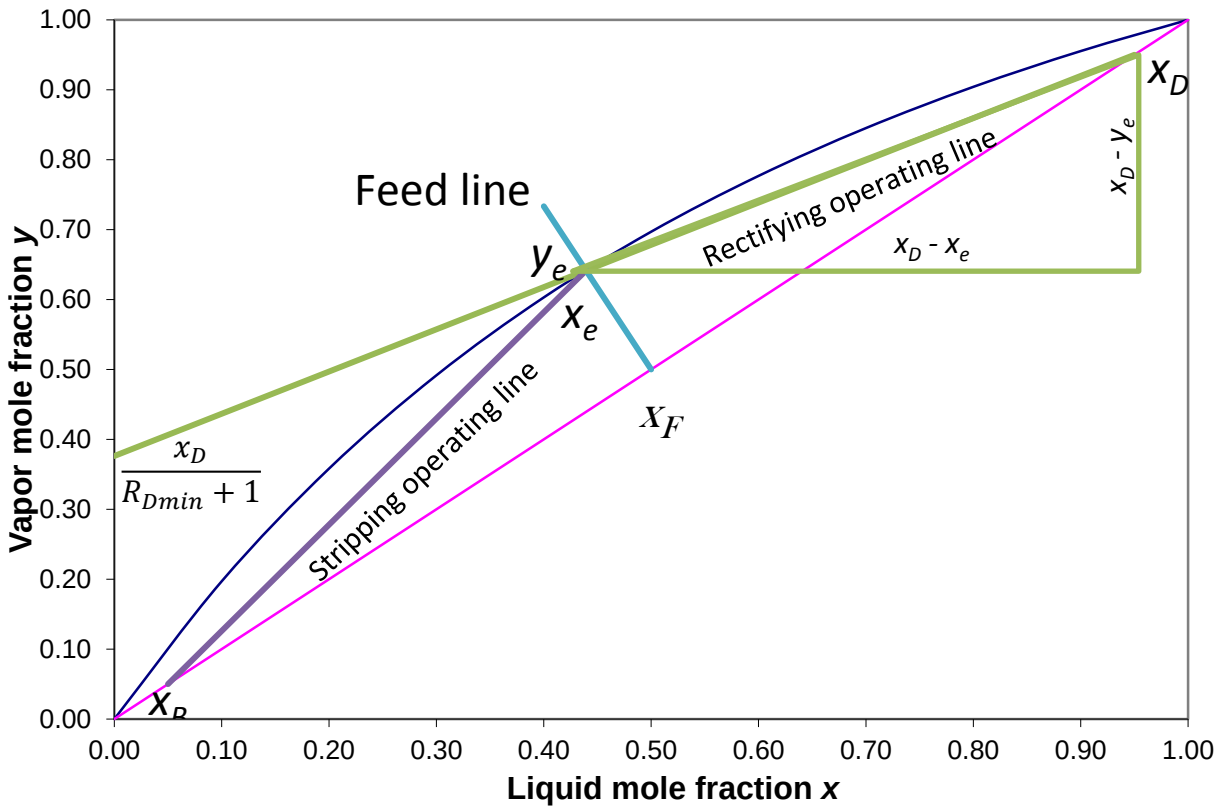


Figure 15. At the minimum reflux ratio the operating lines touch the equilibrium line

There are two ways to determine the minimum reflux ratio from the *x-y* diagram. One is to draw the rectifying operating line that touches the equilibrium line and extend it to the *y*-axis to find the intercept, as in Fig. 15. As before, the intercept is

$$\frac{x_D}{R_{Dmin}+1}$$

Where x_D is the distillate mole fraction and R_{Dmin} is the minimum reflux ratio. Thus the minimum reflux ratio can be determined from the intercept, read from the graph. The other way is to read the coordinates of the point where the feed line crosses

the equilibrium line. By the construction on Fig. 15, the slope of the rectifying operating line is

$$\frac{R_{Dmin}}{R_{Dmin}+1} = \frac{x_D - y_e}{x_D - x_e}$$

Where x_e and y_e are the coordinates of the point where the feed line crosses the equilibrium line. Solving for R_{Dmin}:

$$R_{Dmin} = \frac{x_D - y_e}{y_e - x_e} \tag{24}$$

Optimum Reflux Ratio.

As the reflux ratio is increased from its minimum value, the operating lines separate from the equilibrium line and the required number of trays and the height of the column decrease from infinity, but soon the point of diminishing returns is reached when the number of trays begins to depend more on the pinch points at the top and bottom of the column than on the reflux ratio. As the increase in reflux ratio increases the diameter of the column and the size and heat duties of the condenser and the reboiler, the increased cost in these items overcomes the decrease in the height of the column and the cost of the trays and the total cost of the column increases with reflux ratio. Figure 16 shows a plot of the column total cost versus reflux ratio. For most columns the minimum total cost corresponds to a reflux ratio near 1.2 times the minimum reflux ratio.

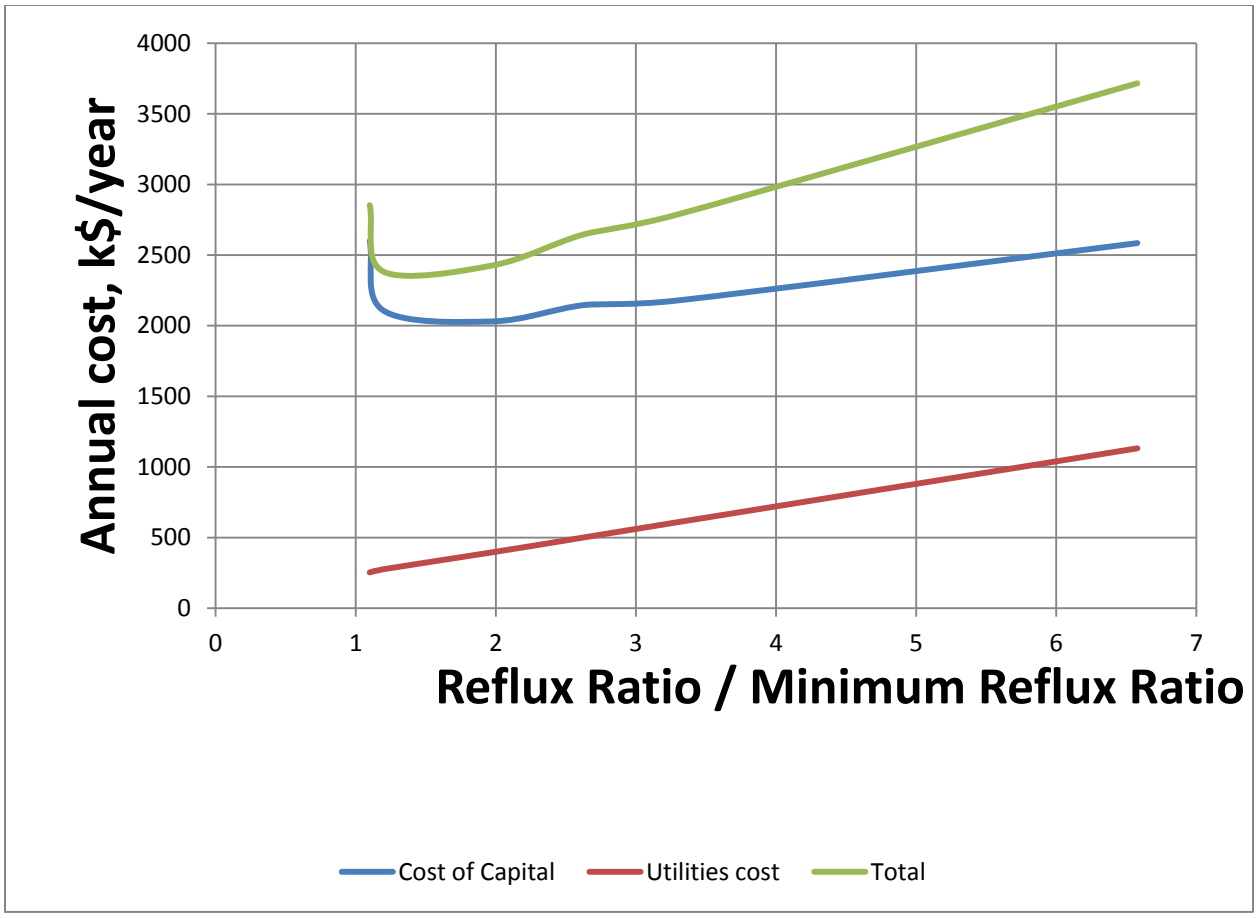


Figure 16. Column annual cost versus reflux ratio

Example 3. Minimum Reflux Ratio for Benzene-Toluene Column.

Determine the minimum reflux ratio and an approximate optimum reflux ratio for the column of Example 2.

Solution. From Fig. 17 we can read that the intercept of the rectifying operating line on the y-axis is 0.373. The minimum reflux ratio is then

$$x_D/(R_{Dmin} + 1) = 0.373$$

$$R_{Dmin} = (0.98/0.373) - 1 = 1.63$$

We can also read the point where the feed line and the equilibrium line cross in Fig. 17: $x_e = 0.45$, $y_e = 0.652$. From Eq. (24) the minimum reflux ratio is,

$$R_{Dmin} = (0.98 - 0.652)/(0.652 - 0.45) = 1.62$$

The values agree within the accuracy of reading the graph. The optimum reflux ratio is near a value of

$$R_D = 1.2R_{D\min} = 1.2(1.62) = 1.95$$

This is the same reflux ratio used in solving Example 2.

Figure 17 also shows that the minimum number of stages, at total reflux, is 8.5.

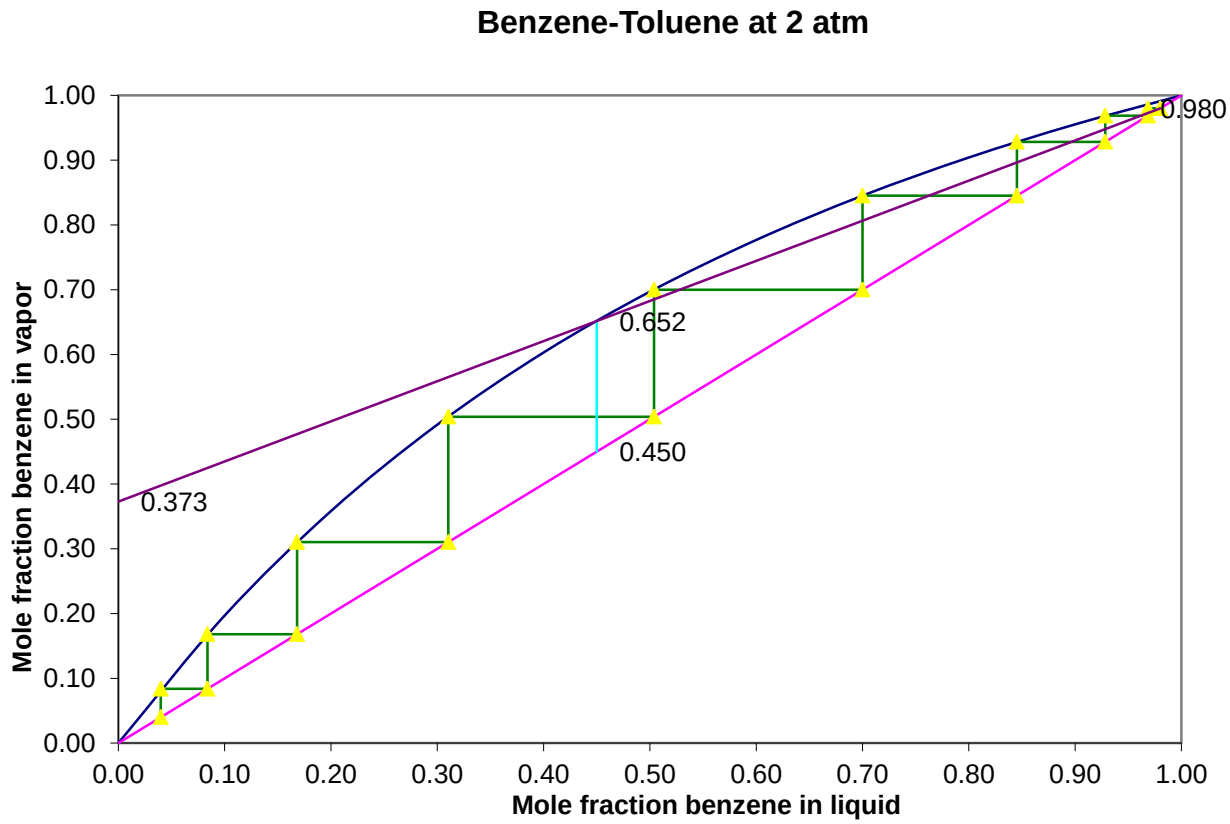


Figure 17. Minimum reflux ratio and minimum number of stages for Example 3

Effect of Feed Enthalpy. As discussed earlier, the feed enthalpy determines the slope of the feed line and, as we saw in the previous discussion, the intersection of the feed line with the equilibrium line determines the minimum reflux ratio and consequently the optimum reflux ratio. Notice from Figs. 15 and 17 that the lower the point where the feed line crosses the equilibrium line, the higher the minimum reflux ratio.

The feed enthalpy can be changed by either heating it or cooling it in a heat exchanger. What would be the advantage of doing so? Figure 7 shows that as the feed goes from sub-cooled liquid superheated vapor, that is, the higher the enthalpy of the feed, the lower the point where the feed line crosses the equilibrium line and thus the higher the reflux ratio. So it would appear to be advantageous to cool the feed as much as possible to reduce the reflux ratio. However, recall from the discussion on feed enthalpy that the sub-cooled feed requires that additional vapor be generated in the reboiler to bring the feed to its bubble point (see Fig. 2a). Therefore, sub-cooling the feed below its bubble point causes an increase in the heat duty of the reboiler and thus its cost.

In conclusion, the feed should enter the column near its bubble point and should be cooled or heated to that condition. However, the economic advantage of pre-heating or cooling the feed to the column is small and should not be considered until the final detailed design of the column. At initial design time it is best to keep the process simple by omitting the column feed exchanger.

Usually the feed to the column may appear to be at its bubble point when it is the product of another column, but such is not the case if the columns operate at different pressures. When the feed comes from a column operating at a higher pressure it will be partially vaporized when it enters the column, and when it comes from a lower-pressure column it will be sub-cooled liquid. In such case a pump must be used to bring the liquid into the higher-pressure column, but pumps are commonly used on liquid feeds to raise the feed to the feed tray which is about half-way up the column. Pumps have a negligible effect on the enthalpy of the liquid feed, and hydrostatic pressure differences do not affect the enthalpy either.

4.3 Effect of Equilibrium.

Separation by distillation requires a difference of volatility between the components being separated. For near ideal solutions one indication of the relative volatility is the difference between the normal boiling points of the components; a convenient indication of how easy or hard a separation by distillation may be is the distance between the equilibrium line and the diagonal line in the *x-y* diagram. This is because, as we have seen, the operating lines fall between the equilibrium line and the diagonal line and the number of required equilibrium stages is higher when the operating lines are closer to the equilibrium line. The reflux ratio is also higher the closer the equilibrium line is to the diagonal line (see Eq. 24).

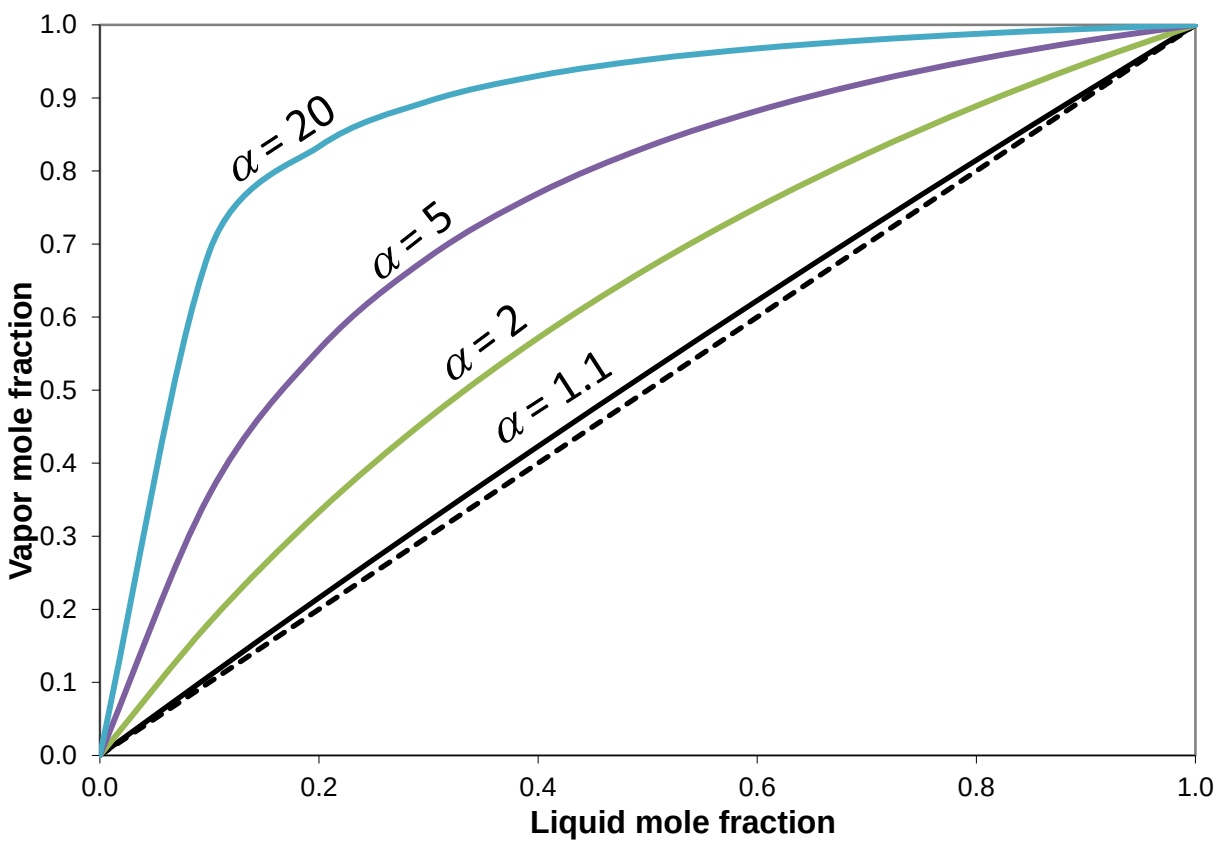


Figure 18. Equilibrium lines for different relative volatilities

Figure 18 shows equilibrium lines calculated at different relative volatilities assuming constant volatilities. We see in the figure that the separation is easier the higher the relative volatility, and harder the closer the relative volatility is to 1.0. The equilibrium line of Fig. 19 for methanol-water shows that the relative volatility is much higher at the bottom of the column than at the top. Because of this it requires more trays at the top of the column to obtain a purer methanol distillate than it does at the bottom to obtain a purer water product. Figure 20 shows the equilibrium line for ethanol-water. This mixture forms an azeotrope at about 87 mole% ethanol (95 volume%) meaning that it is not possible to obtain a distillate product with an ethanol composition higher than 87 mole% in a single binary column. Extractive distillation, involving a third component, is required to separate mixtures that form an azeotrope.

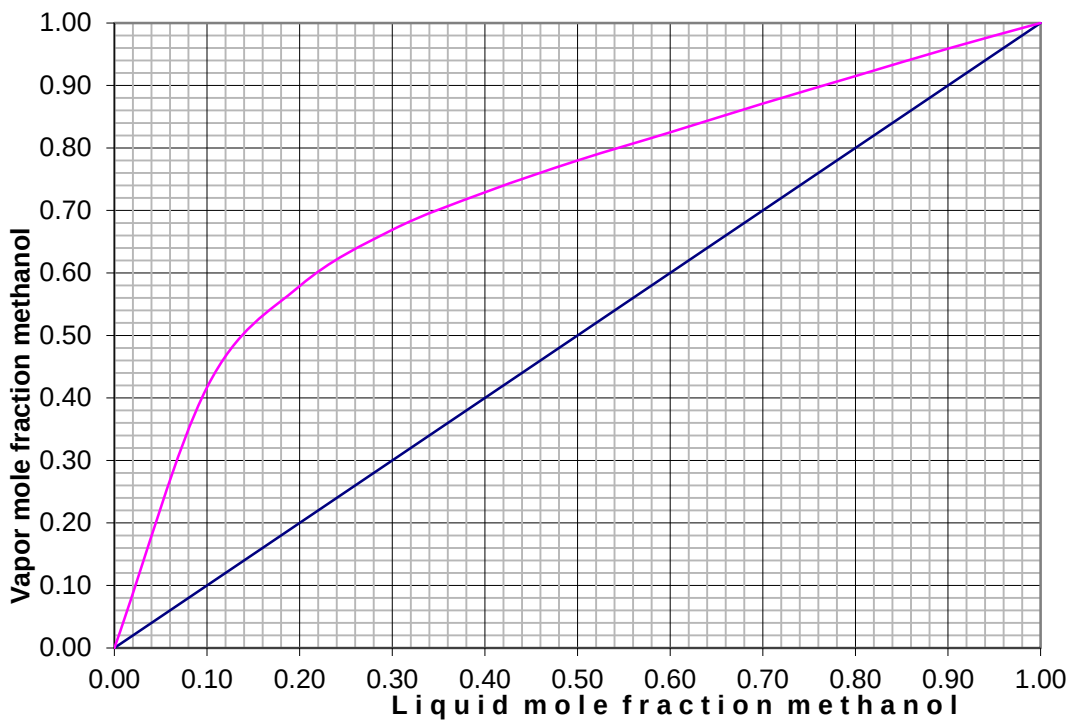


Figure 19. Equilibrium of methanol-water at 1 atmosphere (McCabe, Smith, & Harriott, 7th ed., Table 21.5, page 731)

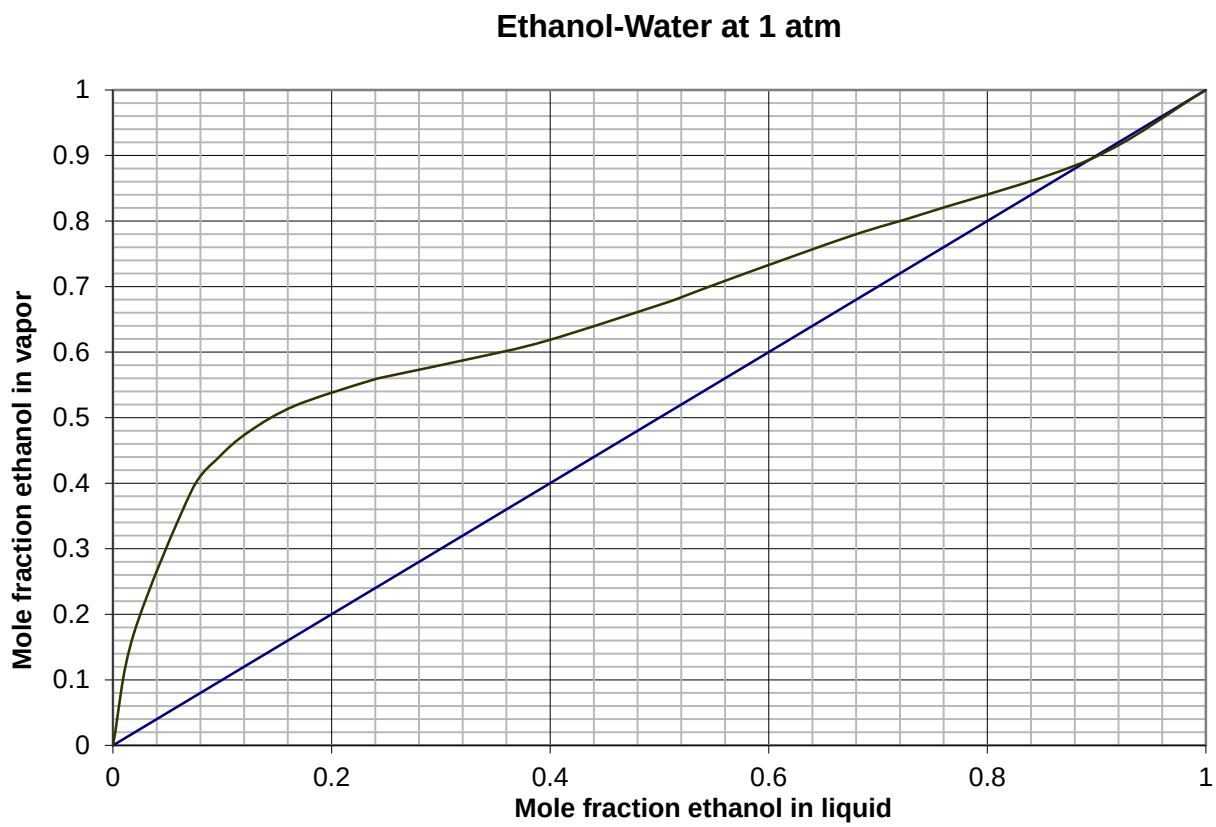


Figure 20. Ethanol-water equilibrium data (Perry's, 7th ed., page 13-12)

In this section we have used the McCabe-Thiele graphical method to understand how the various design specifications affect the size and cost of a column. These insights apply when a simulator is used to do the design calculations, but the consequences of the various decisions are not as easy to see with the simulator as it is with the graphical procedure.

5. Column Sizing

Column tray manufacturers provide their clients with computer programs to do the detailed sizing of distillation columns. This section presents simplified methods to size the columns. Our

objective is to relate the various column design parameters to the resulting height and diameter of the column. We will look at the column efficiency and its effect on the column height, and then at the determination of the column diameter.

5.1 Column Efficiency.

It is common practice, even when using process simulators, to assume the trays on the column are ideal and then use a column efficiency to estimate the required number of actual trays. There are several reasons why column trays are less than ideal:

- There is not enough contact time for the vapor and liquid to reach equilibrium.
- Some of the liquid may leak through the holes in sieve trays to the tray below and by-pass the tray. This phenomena, known as *weeping*, happens at low vapor velocities when the pressure drop across the tray is not sufficient to hold the liquid in the tray.
- At high vapor velocities the vapor may entrain droplets of liquid and carry them into the tray above so that liquid that has already been freed of the heavy component is mixed with liquid that hasn't, a phenomena known as *jet flooding*.

A common measure of the deviation of a tray from ideality is the *Murphree tray efficiency*:

$$\eta_M = \frac{y_k - y_{k+1}}{y_{ek} - y_{k+1}} \tag{25}$$

Where η_M is the Murphree tray efficiency, y_k is the mole fraction of the vapor leaving tray k, and y_{ek} is the vapor mole fraction in equilibrium from the liquid leaving tray k.

Determination of the Murphree tray efficiency as a function of tray parameters is outside the scope of this document, so we will use the *over-all column efficiency*:

$$\eta_o = \frac{N}{N_a} \tag{26}$$

Where η_o is the over-all column efficiency, N is the number of required ideal trays, and N_a is the number of actual trays. Although the number of ideal trays may have a fraction, the number of actual trays must be an integer, because it is not possible to buy and install a fraction of an actual tray. The resulting number from the formula must be rounded to the next higher integer, e.g., 17.3 to 18. In determining the number of required ideal trays we must not include the partial condenser or reboiler when such is present.

5.2 Tower Height.

Once we determine the number of required actual trays from Eq. (26), the height of the tower is estimated by

$$H = N_a \Delta H_s + \Delta H_x \tag{3.31}$$

Where H is the tower height (m), N_a is the actual number of trays, ΔH_s is the tray spacing (m), and ΔH_x is the extra height added to the tray stack for the reflux distributor at the top, the feed distributor at the feed tray, and the space for accumulation of liquid at the bottom of the column (m). The tray spacing or vertical distance between the trays is usually a design premise and is selected in increments of 6 inches (0.1524

m). The extra height is about 15% of the height of the tray stack or 6 m (20 ft), whichever is smaller.

Example 4. Column Height.

Determine the height of the column of Example 2. Assume 65% over-all column efficiency and 24-in tray spacing.

Solution. In Example 2 it was determined that the total number of equilibrium stages is 19. Assuming the usual partial reboiler and total condenser, the number of ideal trays needed is N = 19 -1 = 18.

Number of actual trays: N_a = 18/0.65 = 27.7 or 28 trays

Trays stack height: $N_a \Delta H_s = (28)(24\ in)\frac{1\ m}{39.4 in} = 17.07\ m$

Extra height: ΔH_x = 0.15(17.07 m) = 2.56 m

Total column height: H = 17.07 + 2.56 = 19.6 m (64 ft)

5.2 Column Diameter.

As discussed earlier, when the velocity of the vapors in the column is too high the column efficiency drops due to *jet flooding* or the entrainment of liquid droplets carried and mixed with the liquid in the tray above. Also, at very high vapor velocities the pressure drop across the trays becomes so high that the liquid backs up in the downcomers until the column floods, that is, fills up with liquid. To prevent flooding the column diameter must be such that the vapor velocity is below the *flooding velocity*, that is, the velocity at which the column floods.

Detailed correlations have been developed to estimate the flooding velocity in a column as a function of the liquid and vapor rates and densities, and the liquid surface tension. The surface tension is a measure of the tendency of the liquid to

foam; foaming makes it easier for droplets to be carried by the vapor to the tray above. However, here we present a simple method for estimating the column diameter as a function of the vapor rate, its density, and the tray spacing. Tray manufacturers offer to their clients programs to carry out precise calculations of the trays and the column diameter, but these are outside the scope of this document.

The estimation of the diameter is based on the *F*-factor, defined as follows:

$$F\text{-factor} = v\sqrt{\rho_{Mv} M_w} \tag{28}$$

Where v is the vapor velocity (m/s), ρ_{Mv} is the molar density of the vapors (kgmole/m^3), and M_w is the average molecular weight of the vapors. Values of the *F*-factor at the flooding point are given in Table 1 as a function of the tray spacing (Douglas, 1989). These values are used to estimate the flooding velocity from Eq. (28) and then a smaller velocity is used to estimate the column diameter, e.g., 60% of the flooding velocity.

Table 1. *F*-factor at Flooding

Tray Spacing		*F*-factor at flooding	
inch	m	$\frac{m}{s}\sqrt{\frac{kg}{m^3}}$	$\frac{ft}{s}\sqrt{\frac{lb}{ft^3}}$
12	0.305	1.77	1.45
18	0.457	2.42	1.98
24	0.610	3.06	2.51
36	0.914	3.95	3.24

When the vapors flow up the column not all the column area is available for flow or for bubbling though the liquid in the trays because of the downcomers (see Fig. 6). The downcomers

are where the liquid flows down by gravity from tray to tray. So, the vapor flow is given by

$$V = \rho_{Mv} v \frac{\pi D_i^2}{4}(1 - f_D)\left(\frac{3600\, s}{hr}\right) \tag{29}$$

Where V is the vapor rate (kgmole/hr), D_i is the inside diameter of the column (m), and f_D is the fraction of the column area taken up by the downcomers. Solving for the column diameter

,

$$D_i = \sqrt{\frac{4V}{\pi \rho_{Mv} v (1 - f_D)\left(\frac{3600 s}{hr}\right)}} \tag{30}$$

We see from this formula that the diameter is proportional to the square root of the vapor flow and inversely proportional to the square root of the vapor molar density and of the velocity. As these parameters vary from the top and bottom of the column we must decide which values we should use. We want to estimate the diameter at the point where it is highest because that is the point where the column would flood if the diameter is smaller. The following should be considered:

- When there is large variation of the vapor rate through the column, the point of highest vapor rate must be used to estimate the diameter. The vapor density and the velocity must then be evaluated at that point. If the equimolal overflow assumption I approximately valid, the vapor rate is higher in the rectifying section when the feed is not all liquid as it enters the column, it is higher in the stripping section when the feed is sub-cooled liquid, and is the same in both sections when the feed is saturated liquid.

- If the vapor rate does not vary much, the allowed vapor velocity must be considered. From Eq. (28) we see that the velocity is inversely proportional to the square root of both the vapor molar density and the average molecular weight. So, the higher the molecular weight the lower the flooding velocity and the larger the required column diameter. For most distillation mixtures the high-boiling components have higher molecular weights than the low-boiling components so that the diameter would be larger at the bottom of the column if the vapor rate does not vary much from top to bottom. One exception is water which is often the high-boiling component but has a low molecular weight.
- The molar density of the vapor must also be considered when the vapor rate does not vary much through the column. As the flooding velocity is inversely proportional to the square root of the vapor molar density and the column diameter is inversely proportional to the square root of both the vapor molar density and the velocity, the net effect is that the column diameter is inversely proportional to the fourth root of the vapor molar density which is in turn proportional to the absolute pressure and inversely proportional to the absolute temperature. The temperature is higher at the bottom than at the top of the column, but its effect on the vapor density is small because the relative difference on the absolute temperature from top to bottom is small. The pressure is also higher at the bottom of the column than at the top, but except for columns running under vacuum, the relative difference in pressure from top to bottom is also small.

It follows from the preceding discussion that the diameter must often be determined using the variables at the bottom of the column except when a significant fraction of the feed is vapor as it enters the column, or when the high-boiling component is water.

Estimating the Vapor Molar Density.

For the simplified column sizing method presented here it is acceptable to estimate the vapor molar density assuming the vapors behave as an ideal gas:

$$\rho_{Mv} = \frac{P}{RT} = \left(\frac{1\,\text{kgmole}}{22.4\,\text{m}^3}\right)\left(\frac{273\,\text{K}}{T}\right)\left(\frac{P}{101.3\,\text{kPa}}\right) \tag{31}$$

Where P is the absolute pressure (kPa), T is the absolute temperature (K), and R is the ideal gas law constant (8.31 m^3-kPa/kgmole-K). The value of 22.4 m^3/kgmole is at the standard conditions of 0°C (273 K or 492°R) and atmospheric pressure (101.3 kPa or 14.7 psia).

When the results of the column simulation are used to size the column, most simulators provide a column profile showing the vapor and liquid rates, and the temperature, pressure and average molecular weight on each tray. However, when the column is designed by the McCabe-Thiele graphical procedure, the temperature and average molecular weight are not usually parts of the column specifications, requiring they be estimated. To determine the diameter at the bottom of the column we use the composition of the bottoms product to estimate its average molecular weight and its bubble point to be used as the temperature in estimating the molar density. For the diameter at the top of the column the average molecular weight and dew point

of the distillate are used, as the vapors leave the top of the column at the composition of the distillate.

Example 5. Column Diameter Estimation.

Estimate the diameter of the benzene-toluene column in of Example 2 using 60% of the flooding velocity and allow 12% of the column area for the downcomers. The tray spacing is 24 in.

Solution. In Example 2 we determined that the vapor rate in the column is 708 kgmole/hr, the same at the top and bottom of the column. The pressure is also uniform at 2 atm. We will estimate the diameter at the bottom as toluene has a higher molecular weight. The composition of the bottoms product is 4 mole% benzene.

Average molecular weight: 0.04(78) + 0.96(92) = 91.4 kg/kgmole

Dew point of bottoms product (use Antoine constants (Reid, Prausnitz, and Sherwood, 1977, Appendix) :

Benzene: $ln\left(\frac{P_B^o}{mm\,Hg}\right) = 15.9008 - \frac{2788.51K}{T-52.36K}$

Toluene: $ln\left(\frac{P_T^o}{mm\,Hg}\right) = 16.0137 - \frac{3096.52K}{T-53.67K}$

Bubble point (assume Raoult's Law applies):

$$0.04P_B^o + 0.96P_T^o = (2\,atm)\left(\frac{760\,mm\,Hg}{atm}\right)$$

Solving iteratively, we find: T = 408 K

Vapor molar density: $\rho_{Mv} = \left(\frac{kgmole}{22.4\,m^3}\right)\left(\frac{273K}{408K}\right)\left(\frac{2\,atm}{1\,atm}\right) = 0.0597\,\frac{kgmole}{m^3}$

Velocity (use F-factor from Table 1 and 60% of flooding velocity): $v = 0.60\left(\frac{3.06\,m}{s}\sqrt{\frac{kg}{m^3}}\right)\sqrt{\left(\frac{m^3}{0.0597\,kgmole}\right)\left(\frac{kgmole}{91.4\,kg}\right)} = 0.786\,\frac{m}{s}$

Column diameter (allowing 12% of the column area for the downcomers):

$$D_i = \sqrt{\frac{4}{\pi}\left(\frac{708\,kgmole}{hr}\right)\left(\frac{s}{0.786\,m}\right)\left(\frac{m^3}{0.0597\,kgmole}\right)\left(\frac{hr}{3600\,s}\right)\left(\frac{1}{1-0.12}\right)} = 2.5\,m$$

So the diameter must be 2.5 m or 8.5 ft.

As we shall see later, the column diameter affects the cost of the column and the cost of the trays. Next we will look at the sizing of the condenser and the reboiler.

6. Condenser and Reboiler Sizing

The detailed design of heat exchangers was the subject of earlier lectures. The complete design of heat exchangers requires correlations to estimate the coefficients of heat transfer and the selection of the shell diameter and the number, diameter and length of tubes. Here we will present only the approximate estimation of the required heat transfer area of the condenser and the reboiler. This is usually the first step in the full detailed design of the exchangers.

6.1 Condenser Area.

To estimate the heat transfer area of the condenser we will assume no sub-cooling of the condensate so that the heat duty is only that required to condense the vapors. We will also assume the vapors condense at a uniform temperature, the bubble point of the distillate in a total condenser or the dew point of the vapor distillate in a partial condenser.

For a total condenser all of the vapors leaving the top of the column must be condensed (see Fig. 9), while on a partial condenser with a single vapor distillate product only the portion of the vapors that return to the column as reflux (*full reflux*) are condensed (see Fig. 9). The heat duty is then estimated by,

Total condenser: $$Q_c = V_R \lambda_c \qquad (32)$$

Partial condenser, full reflux: $Q_c = L_R\lambda_c$ (33)

Where Q_c is the condenser heat duty (kJ/hr), λ_c is the latent heat of the vapors (kJ/kgmole), and V_R and L_R are the rectifying vapor and reflux flows (kgmole/hr), respectively. When a simulator is used to design the column, the simulator accurately calculates the condenser heat duty and Eqs. (32) and (33) are not necessary.

The temperature difference in the condenser depends on the cooling medium used.

- If an air condenser is used, the exchanger is a finned cross-flow exchanger and it is usually assumed the air temperature is just the ambient temperature T_{air}. Then the temperature difference is $\Delta T_c = T_c - T_{\text{air}}$.
- When refrigerant must be used due to the low condenser temperature, the refrigerant is assumed to vaporize at uniform temperature T_{ref} and the temperature difference is $\Delta T_c = T_c - T_{\text{ref}}$.
- When the cooling medium is cooling tower water, the temperature of the water varies from inlet to outlet, as Fig. 21 shows. Then the logarithmic mean temperature difference must be used:

$$\Delta T_c = \frac{(T_c - T_{wi}) - (T_c - T_{wo})}{ln\left(\frac{T_c - T_{wi}}{T_c - T_{wo}}\right)} \quad (34)$$

Where ΔT_c is the mean temperature difference (°C), T_c is the condenser temperature (°C), and T_{wi} and T_{wo} are the inlet and outlet temperatures of the coolant (°C), respectively.

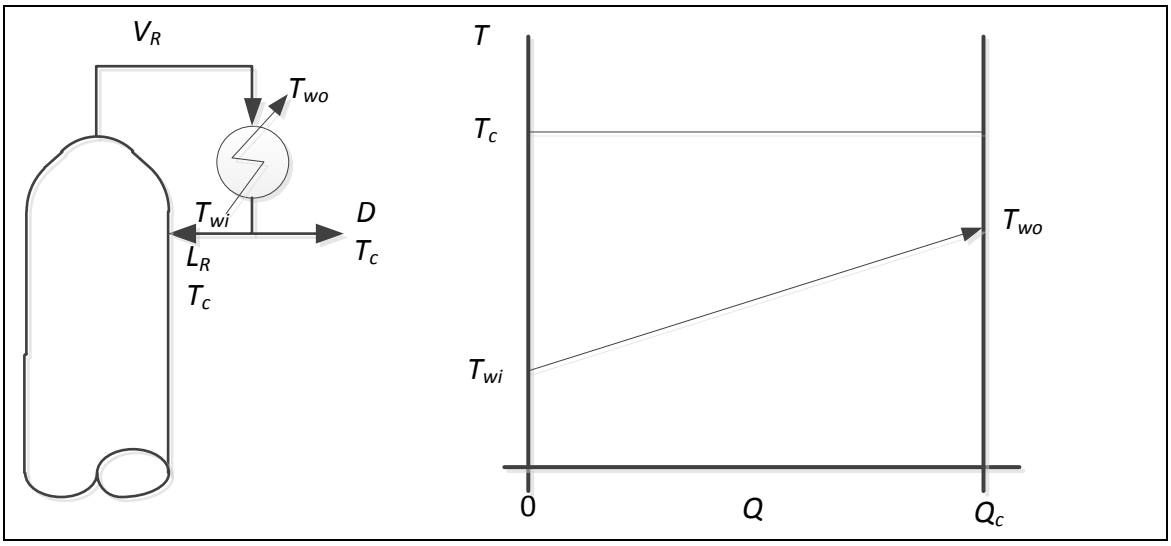


Figure 21. Total condenser schematic and temperature profile

Condenser Restrictions.

When selecting the temperatures for the condenser engineers are restricted by certain design premises. The inlet cooling water temperature is set by the ambient conditions at the plant location, usually the summer conditions. The temperature rise of the water through the exchanger is restricted to about 10°C (20°F) because high return temperatures may foul the cooling tower. Finally, a minimum temperature approach (difference) of about 5°C (10°F) is allowed to prevent excessively large exchanger areas. So, assuming an inlet cooling water temperature of 30°C (80°F), the maximum outlet temperature is 40°C (100°F) (the maximum temperature rise is selected to avoid requiring unnecessarily large cooling water flow). We see then that for a minimum temperature approach of 5°C (10°F) the minimum condenser temperature is 45°C (110°F). As discussed earlier, this minimum condenser temperature required to avoid using refrigerant is a major consideration in selecting the column pressure, because refrigerant is much more expensive than cooling water.

The last parameter required to estimate the required condenser area is the over-all heat transfer coefficient. As a detailed estimation is not possible until a specific exchanger is selected, rough estimates are used to get a preliminary estimate of the area. Table 2 shows some typical coefficients. We will use these as estimates in our design procedure.

Table 1. Typical Condenser Coefficients

Cooling medium	Condensing vapors	U_{oc}, kJ/hr-m^2-°C	U_{oc}, BTU/hr-ft^2-°F
Air	Any	200	10
Cooling Water	Aromatics	1,400	70
	Light hydrocarbons	1,800	90
	Chlorinated hydrocarbons	1,800	90
Refrigerant	Aromatics	1,800	90
	Light hydrocarbons	2,400	120
	Chlorinated hydrocarbons	2,400	120

The condenser area is then estimated from the definition of the over-all heat transfer coefficient:

$$A_c = \frac{Q_c}{U_{oc}\Delta T_c} \tag{35}$$

Where A_c is the condenser area (m^2) and U_{oc} is the estimated over-all heat transfer coefficient (kJ/hr-m^2-°C).

The required flow of cooling water in kg/hr is given by $Q_c/[c_{pw}(T_{wo} - T_{wi})]$ where c_{pw} is the specific heat of water (4.18 kJ/kg-°C or 1.0 Btu/lb-°F).

Example 6. Condenser Sizing.

Size the condenser for the benzene-toluene column of Example 2. Assume the coolant is cooling tower water entering at 30°C with a temperature rise of 10°C. The minimum approach is 5°C. Determine also the required flow of cooling water.

Solution. To estimate the heat duty and temperature of the condenser we need the average latent heat of the vapor and the bubble point of the distillate. From Example 2 we know the distillate composition is 98 mole% benzene and 2 mole% toluene. Once again we assume Raoult's law applies and use the Antoine equations of Example 5 to find the vapor pressures and the bubble point of the condensate:

Raoult's law: $0.98P_B^o + 0.02P_T^o = 2\ atm\left(\frac{760\ mmHg}{atm}\right)$

Solve simultaneously to obtain the bubble point: T_c = 378 K

From Perry's, 8th ed., Table 2-150, we find the latent heats of benzene and toluene (critical temperatures of 562 K for benzene and 592 K for toluene from Table 2-141):

Benzene: $T_{rB} = \frac{378\ K}{562\ K} = 0.673 \quad \lambda_B = 4.5346 \times 10^7(1-0.673)^{0.39053} = 29300\ \frac{kJ}{kgmole}$

Toluene: $T_{rT} = \frac{378\ K}{592\ K} = 0.639 \quad \lambda_T = 4.9507 \times 10^7(1-0.639)^{0.37742} = 33700\ \frac{kJ}{kgmole}$

Average latent heat: $\lambda_c = 0.98(29300) + 0.02(33700) = 29{,}400\ \frac{kJ}{kgmole}$

The vapor rate in the rectifying section is, from Example 2, 708 kgmole/hr. The condenser temperature is 378 K – 273 K = 105°C.

Heat duty: $Q_c = \left(708\frac{kgmole}{hr}\right)\left(29{,}400\frac{kJ}{kgmole}\right) = 20.8 \times 10^6\ \frac{kJ}{hr}$

Mean temperature difference: $\Delta T_c = \frac{(105-30)-(105-40)}{ln\left(\frac{105-30}{105-40}\right)} = 70°C$

From Table 3.2, estimate the heat transfer coefficient as U_{oc} = 1,400 kJ/hr-m^2-°C.

Condenser area: $A_c = \left(\frac{20.8\times10^6\ kJ}{hr}\right)\left(\frac{hr-m^2-°C}{1{,}400\ kJ}\right)\frac{1}{70°C} = 212\ m^2$

The condenser must have 212 m^2 or 2,200 ft^2.

Cooling water flow: $\left(\frac{20.8\times10^6\ kJ}{hr}\right)\left(\frac{kg-°C}{4.18kJ}\right)\left(\frac{1}{40°C-30°C}\right) = 498{,}000\ \frac{kg}{hr}$

Or 8,300 liters/min (2,200 gallons/minute).

6.2 Reboiler Sizing.

The heating medium for reboilers is usually steam, except when the reboiler temperature is too high for the available steam pressures which case the reboiler must be a furnace heated by a burning fuel. In some cases heating oil is heated in a furnace and circulated to the reboiler.

When steam is used as the heating medium it condenses at uniform temperature T_s outside the reboiler tubes and the liquid boils at a uniform temperature T_R, assumed to be the bubble point of the bottoms product.

The temperature difference is then $\Delta T_R = T_s - T_R$.

As condensing steam has one of the lowest resistances to heat transfer, the over-all heat transfer coefficient of the reboiler depends on the resistance of the boiling liquid. This resistance is highly dependent on the boiling regime which is in turn dependent on the temperature difference. At low temperature differences, less than 20°C (35°F), heat is transferred by natural convection and the heat transfer coefficient is very low. When the temperature difference is between 20 and 25°C (35 to 45°F), the liquid boils by rapid generation of bubbles, a regime known as *nucleate boiling* which is very efficient for the transfer of heat. The heat transfer coefficient is around 5,000 kJ/hr-m^2-°C (250 BTU/hr-ft^2-°F). At higher temperature differences the liquid boils so rapidly that a film of vapor forms around the tubes insulating them and causing the heat transfer coefficient to drop rapidly. This regime, known as *film boiling*, is to be avoided. So, the design of a reboiler proceeds as follows:

- Select a steam pressure that provides at least a 20°C (35°F) temperature difference to insure nucleate boiling.

- If the steam pressure available provides a temperature difference greater than 25°C (45°F), drop the steam pressure with a valve to make the temperature difference 25°C. This avoids film boiling.
- Assume a heat transfer coefficient of 5,000 kJ/hr-m^2-°C (250 BTU/hr-ft^2-°F.

When the reboiler is a furnace it is not necessary to determine its area, because furnaces are sized on the basis of their heat duty.

The heat duty of the reboiler is the heat required to vaporize the vapors:

$$Q_R = V_S \lambda_R \tag{36}$$

Where Q_R is the reboiler heat duty (kJ/hr), V_S is the flow of vapors into the column (kgmole/hr), and λ_R is the average latent heat of vaporization (kJ/kgmole) at the bottoms product composition and bubble point temperature.

The reboiler area is then $A_R = \frac{Q_R}{U_{oR} \Delta T_R}$

where U_{oR} is the over-all heat transfer coefficient (kJ/hr-m^2-°C) and ΔT_R is the temperature difference (°C).

The required flow of steam in kg/hr is given by Q_R/λ_s, where λ_s is the latent heat of the steam (kJ/kg).

Example 7. Reboiler Sizing.

Size the reboiler for the benzene-toluene column of Example 2. Determine also the required steam pressure to insure nucleate boiling and the required flow of steam.

Solution. It was determined in Example 2 that the flow of vapors is 708 kgmole/hr and the composition of the bottoms product is 4

mole% benzene, and in Example 5 that the bubble point of the bottoms product is 408 K – 273 K = 135°C.

Determine the average latent heat of vaporization of the reboiler liquid, as in Example 6:

Benzene: $T_{rB} = \frac{408\,K}{562\,K} = 0.726 \quad \lambda_B = 4.5346 \times 10^7 (1 - 0.726)^{0.39053} = 27400\ \frac{kJ}{kgmole}$

Toluene: $T_{rT} = \frac{408\,K}{592\,K} = 0.689 \quad \lambda_T = 4.9507 \times 10^7 (1 - 0.689)^{0.37742} = 31900\ \frac{kJ}{kgmole}$

Average latent heat: $\lambda_c = 0.04(27400) + 0.96(31900) = 31{,}700\ \frac{kJ}{kgmole}$

Reboiler heat duty: $Q_R = \left(708\frac{kgmole}{hr}\right)\left(31{,}700\frac{kJ}{kgmole}\right) = 22.5 \times 10^6\frac{kJ}{hr}$

For nucleate boiling the steam temperature must be at least 135°C + 20°C = 155°C. At this temperature the steam pressure is 543 kPa (79 psia), and its latent heat of vaporization is 2099 kJ/kg (Keenan, Keyes, Hill and Moore, *Steam Tables*, 1978.

Reboiler area: $A_R = \left(22.5 \times 10^6\frac{kJ}{hr}\right)\left(\frac{hr-m^2-°C}{5{,}000\,kJ}\right)\left(\frac{1}{20°C}\right) = 225\,m^2$

Or 2,400 ft^2.

Required steam flow: $\frac{Q_R}{\lambda_s} = \left(22.5 \times 10^6\frac{kJ}{hr}\right)\left(\frac{kg}{2099\,kJ}\right) = 10{,}700\frac{kg}{hr}$

This section has presented a simple procedure for sizing distillation columns, that is, for determining their height, diameter, and the areas of the condenser and the reboiler, as well as the consumption of cooling and heating utilities in the condenser and reboiler, respectively. The next section presents correlations for estimating the cost of the column from these sizing parameters.

7. Column Cost

Chemical engineers usually depend on expert cost estimators in their organizations to estimate the precise cost of equipment.

Because of this they don't usually look at equipment costs until the final design is done. However, it is convenient to have some approximate cost correlations to be able to evaluate design decisions without having to refer to the cost estimator every time a decision is made. Although these correlations are not accurate they give approximate order-of-magnitude costs that are good enough to evaluate the design decisions.

The following correlations have been adapted from those given by Douglas (1989). They give estimates of the installed cost of equipment, including structures, piping, insulation, instrumentation, foundations and painting. One attractive advantage is that the factor for installation costs is separate from the cost factor for material of construction, pressure rating, and equipment type. These cost factors can be high but have little effect on the installation costs. Also the installation cost factors are different for each type of equipment, being smaller for such high-cost equipment as furnaces than for lower-cost equipment like vessels and exchangers. This does not mean that the installation of high-cost equipment is lower than for low-cost equipment, just that it is not proportional to the equipment cost.

The column cost is estimated using the correlation for pressure vessels and is based on the height and diameter of the column:

$$IC_{col} = \left(\frac{M\&S}{280}\right)\$940\left(\frac{D_i}{m}\right)^{1.066}\left(\frac{H_c}{m}\right)^{0.802}(F_c + 2.18) \qquad (37)$$

Where IC_{col} is the installed cost of the column ($), M&S is the Marshall and Swift inflation index, published each month in *Chemical Engineering* magazine, D_i and H_c are the column diameter and height (m), respectively, and F_c is a cost factor that

depends on the column design pressure and the material of construction. The correlation is good for diameters up to 10 m (30 ft) and heights up to 120 m (400 ft). The factor 2.18 is the one that accounts for installation costs.

When the column is made in carbon steel—handling non-corrosive chemicals—and for operating pressure below 345 kPa gage (50 psig), the cost factor F_c is unity. For higher pressures and other materials of construction, the factor is:

$$F_c = F_p + F_m$$

Where the pressure and material factors are

For design pressures up to

kPa g	345	690	1380	2070	2760	3450	4140	4820	5520	6200	6900
psig	50	100	200	300	400	500	600	700	800	900	1000
F_p	0	0.05	0.15	0.20	0.35	0.45	0.60	0.80	0.90	1.30	1.50

Factor for material of construction:

Material	Carbon steel	Stainless steel	Monel	Titanium
F_m, clad	1.00	2.25	3.89	4.25
F_m, solid	1.00	3.67	6.34	7.89

Where "clad" means that the column is made of carbon steel with a protective inner layer of the specified material clad to it, and "solid" means that the entire column is made of the specified material.

The cost of the trays is

$$IC_{trays} = \left(\frac{M\&S}{280}\right) \$60 \left(\frac{D_i}{m}\right)^{1.55} N_a F_c \qquad (38)$$

Where IC_{trays} is the installed cost of the trays ($), D_i is the column diameter (m), N_a is the number of actual trays and F_c is a

cost factor which is 1.0 for sieve trays in carbon steel and 24-in tray spacing. For other conditions it is given by

$$F_c = F_s + F_t + F_m$$

Where F_s is

- for 24-in tray-spacing,
- 1.05 for 18-in spacing and
- 1.10 for 12-in spacing,

F_t is

- 0 for sieve trays,
- 0.4 for valve trays, and
- 1.8 for bubble-cap trays,

and F_m is

- 0 for carbon steel,
- 1.7 for stainless steel, and
- 8.9 for monel.

The cost of the condenser and reboiler can be estimated by the following correlation:

$$IC_{exch} = \left(\frac{M\&S}{280}\right) \$480 \left(\frac{A}{m^2}\right)^{0.65} (F_c + 2.29) \tag{39}$$

Where IC_{exch} is the installed cost of the exchanger ($), A is the heat transfer area (m^2), and F_c is a cost factor which is 1.0 for a floating head exchanger in carbon steel with a design pressure of less than 1000 kPa gage (150 psig). The correlation is valid for exchangers of up to 460 m^2 (5,000 ft^2). When the area is larger than the correlation limit, the cost is best estimated by assuming more than on exchanger all with the same area less than

the limit, e.g., for a 900 m^2 exchanger assume three exchangers of 300 m^2 each.

For other exchanger types, design pressures and materials of construction, the cost factor is

$$F_c = F_m(F_d + F_p)$$

Most condensers are of the floating head type for which F_d = 1.00, or *U*-tube type (F_d = 0.85), and most modern reboilers are of the fixed head type—thermo-syphon—for which F_d = 0.80. The factors for material of construction and the design pressure are:

Cost factors for material of construction, shell/tube

Material	CS/CS	CS/Brass	CS/MO	CS/SS	SS/SS	CS/Monel	Monel/Monel	CS/Ti	Ti/Ti
F_m	1.00	1.30	2.15	2.81	3.75	3.10	4.25	8.95	13.05

Cost factors for design pressures up to

kPa gage	1030	2070	2760	5510	6900
psig	150	300	400	800	1000
F_p	0.00	0.10	0.25	0.52	0.55

One important point to keep in mind when using these cost correlations is that their precision is limited to about two significant digits, that is, they are only good to make rough order-of-magnitude estimates of the cost. It would be very wrong to report the cost results to more than three significant digits.

Example 8. Column Cost Estimation.

Estimate the cost of the benzene-toluene column of Example 2.

Solution. The column requires 28 actual trays with a height of 19.6 m (Example 4) and a diameter of 2.5 m (Example 5). As

benzene and toluene are not corrosive the column can be made of carbon steel and for the operating pressure less than 345 kPa gage, the cost factor is 1.0. Also, for sieve trays in carbon steel with 24-in tray spacing, the cost factor for the trays is also 1.0. Let's use an M&S inflation index of 1600 (around 2010 costs):

Column Cost: $IC_{col} = \left(\frac{1600}{280}\right)\$940\left(\frac{2.5m}{m}\right)^{1.066}\left(\frac{19.6m}{m}\right)^{0.802}(1.0 + 2.18) = \$490 \times 10^3$

Tray Cost: $IC_{trays} = \left(\frac{1600}{280}\right)\$60\left(\frac{2.5m}{m}\right)^{1.55} 28(1.0) = \$40 \times 10^3$

The condenser area is 212 m^2 (Example 6). For a floating-head exchanger in carbon steel at a pressure below 1000 kPa gage the cost factor is 1.0.

Condenser Cost: $IC_c = \left(\frac{1600}{280}\right)\$480\left(\frac{212m^2}{m^2}\right)^{0.65}(1.0 + 2.29) = \$290 \times 10^3$

The reboiler area is 225 m^2 (Example 7). For a thermosyphon reboiler (fixed tube) in carbon steel at a pressure below 1000 kPa gage the cost factor is 0.8.

Reboiler Cost: $IC_R = \left(\frac{1600}{280}\right)\$480\left(\frac{225m^2}{m^2}\right)^{0.65}(0.8 + 2.29) = \$290 \times 10^3$

The total installed cost of the column is then approximately $(490+40+290+290)\times10^3$ = \$1.10 million.

Although these correlations are not very precise, they serve to study the effect of design parameters such as the reflux ratio on the total cost of the column, and to compare the costs of the various columns in a process. For example, if in designing a process with the column in the preceding example we find there is another column that costs \$5 million, it would be better to concentrate in reducing the cost of the more expensive column.

When considering the total cost of the column it is important to consider also the annual cost of utilities for the

condenser and the reboiler. Although the cost of cooling water is small—less than a dollar per million kJ—that of refrigeration is much higher—over $10 per million kJ. Reboiler steam can cost from $5 to $12 per million kJ, depending on its pressure (200 to 4000 kPa).

8. Column Simulation

Most simulators provide two different models for distillation columns, a sometimes called "short-cut" model and a rigorous model. Unfortunately, the rigorous model requires the specification of the number of trays and the feed tray location, parameters that are not usually available when the column is being designed. The short-cut model, similar to the McCabe-Thiele graphical method provides the number of required equilibrium stages for a specified reflux ratio and separation specifications, but it is not as accurate as the rigorous model. So, a good way to simulate a non-existing column is to first use the short-cut method to determine the number of stages—some simulators also provide an estimate of the feed tray location—and then use the results of the short-cut model to run the rigorous model of the column. When doing this it is important to follow the following guidelines:

- The pressures specified for the short-cut model must match exactly the pressures to be specified in the rigorous model. Therefore the condenser and reboiler temperatures must be checked when the pressures are specified in the short-cut model so that they do not have to be changed in the rigorous model.
- The separation specifications—product compositions—in the short-cut model must match the corresponding specifications in

the rigorous model because, just as with the pressures, the results of the short-cut model depend on these specifications.

- The reflux ratio in the short-cut model should be specified near 1.2 times the minimum reflux ratio because, as we have seen, it is usually near the optimum reflux ratio. The short-cut model provides the value of the minimum reflux ratio, while the rigorous model does not. This is because for the specifications required by the rigorous model the minimum reflux ratio has no meaning.

The number of trays, feed tray location and reflux ratio obtained from the short-cut model are then specified in the rigorous model, along with the column pressures and the flow of the distillate. After the rigorous model converges, because of the different manner of doing the calculations, the separation may be different than for the short-cut model. Then the reflux ratio and distillate flow specifications are replaced with the correct separation specifications—product compositions or recoveries—so that the rigorous model varies the reflux ratio to achieve the correct separation. In our experience this procedure results in an acceptable column design with the best changes for convergence of the rigorous model.

You may ask, if the rigorous model results in a different reflux ratio than the short-cut model, how do we know that the reflux ratio is still near optimum? In fact we don't, except that the number of trays obtained from the short-cut model should still be near optimum provided that the reflux ratio is not much different. This procedure should be better than trying different number of trays in the rigorous model looking for an optimum, for this would require doing a full optimization study based on the total cost of the column and its utilities. Engineers seldom have time for such a study when doing the preliminary design of a process.

Consider also this: as we have seen in this section, the number of required equilibrium stages depends not only on the reflux ratio but also on the separation specifications, the pressure-dependent equilibrium relationships of the components, and the feed composition and enthalpy. The short-cut model takes all these parameters into consideration when estimating the required number of stages, while the engineer picking the number of trays out of the air doesn't.

Summary

These lecture notes have presented the design of binary distillation columns using the McCabe-Thiele graphical procedure as a basis for discussion. The objective has been to get an insight on how the various column design parameters affect the size and cost of the column. Such an insight should be invaluable in guiding the design of distillation columns using process simulators.

The McCabe-Thiele method showed us how to determine the minimum and near optimum reflux ratio, the required number of equilibrium stages and the feed tray location. Simple methods were presented to estimate the column height and diameter and the heat duties and areas of the condenser and the reboiler. Correlations were also presented to estimate the installed costs of the column, the trays, the condenser and the reboiler.

Review Questions

1. What property of the components in a mixture allows them to be separated by distillation?

2. Name the product streams of a distillation column? To which of the products does the bulk of the *light* or more volatile component go?
3. How many degrees of freedom are there in the design of a binary distillation column?
4. How many feed variables are specified in a design problem for a binary distillation column?
5. Other than the feed variables, which other specifications are necessary to design a distillation column?
6. Which is the *rectifying section* of a column? Which is the *stripping section*?
7. What is the purpose of the *condenser*? What is the purpose of the *reboiler*?
8. Define the reflux ratio.
9. When the reflux ratio is decreased, how does the required number of equilibrium stages change?
10. When the purity of the products is increased, how does the number of required equilibrium stages change?
11. When the relative volatility of the components is higher, how does the number of required equilibrium stages change?
12. When the feed rate to a distillation column is increased, how does the number of required equilibrium stages change?
13. What is total reflux? What is the number of required equilibrium stages at total reflux?
14. What is minimum reflux? What is the number of required equilibrium stages at total reflux?
15. What is *weeping*? What is *jet flooding*? When do each of these occur in terms of the vapor velocity?
16. How does the diameter of the column change when the feed rate to the column is higher? How does it change when the reflux ratio is increased?

17. How do the heat duties and areas of the condenser and the reboiler change when the feed rate to the column is higher? How do they change when the reflux ratio is increased?
18. Which is the main consideration in selecting the operating pressure of a column?
19. When simulating a column, why is it important to use both the short-cut model and the rigorous model?
20. Why must the specifications of the column pressures and product purities of the rigorous model match those of the short-cut model?

Problems

1. ***Distillation Column Material Balances.*** The feed to a distillation column consists of 600 kmole/hr of a binary mixture with a composition of 35 mole% of the light component. Draw a schematic of the column showing all the problem data and determine:
   a. The distillate and bottoms product rates and component recoveries when the distillate composition is 0.5 mole% of the heavy component and the bottoms composition is 0.1 mole% of the light component.
   b. The distillate and bottoms product rates and bottoms composition when 98% of the light component is recovered in the distillate with composition of 0.5 mole% of the heavy component.
   c. The maximum possible composition of the distillate if the distillate rate is fixed at 250 kmole/hr.
   d. The maximum recovery of the light component if the bottoms rate is fixed at 400 kmole/hr.

2. **Distillation Column Material Balances.** A distillation column is fed 1,200 kmole/hr of a mixture containing 60 mole% component A and 40 mole% component B, where A is the more volatile component. Draw a schematic of the column showing the problem data and determine:
   a. The distillate and bottoms rates if the distillate is 1 mole% B and the bottoms is 5 mole% A
   b. The distillate and bottoms rates for 99% recovery of A and a distillate purity of 98 mole% A
   c. The maximum mole fraction of A in the distillate when the distillate rate is 800 kmole/hr
   d. The minimum mole fraction of A in the bottoms when the distillate rate is 650 kmole/hr.

3. **Balances on Methanol-Water Column.** A saturated liquid containing 55 weight% methanol and the balance water is fed to a distillation column at the rate of 4,000 kg/min. Draw a schematic of the column showing the problem data and determine:
   a. The rate of the distillate and bottoms products and the bottoms composition when 99.5% of the methanol is recovered and the composition of the bottoms is 1.0 weight% methanol.
   b. The maximum composition of methanol in the distillate when the distillate rate is 5,400 kmole/hr
   c. The rates of the distillate and bottoms products and component recoveries when the distillate contains 3 weight% water and the bottoms contains 1.0 weight% methanol.

4. ***Design of Methanol-Water Column by McCabe-Thiele.*** For the data of Problem 3, part c, draw the schematic of the column

showing the problem data and determine the number of required equilibrium stages and the feed tray location at a reasonable reflux ratio. Report also the vapor and liquid rates in the rectifying and stripping sections of the column. Equilibrium data at 1 atm are given in Fig. 19 of the notes.

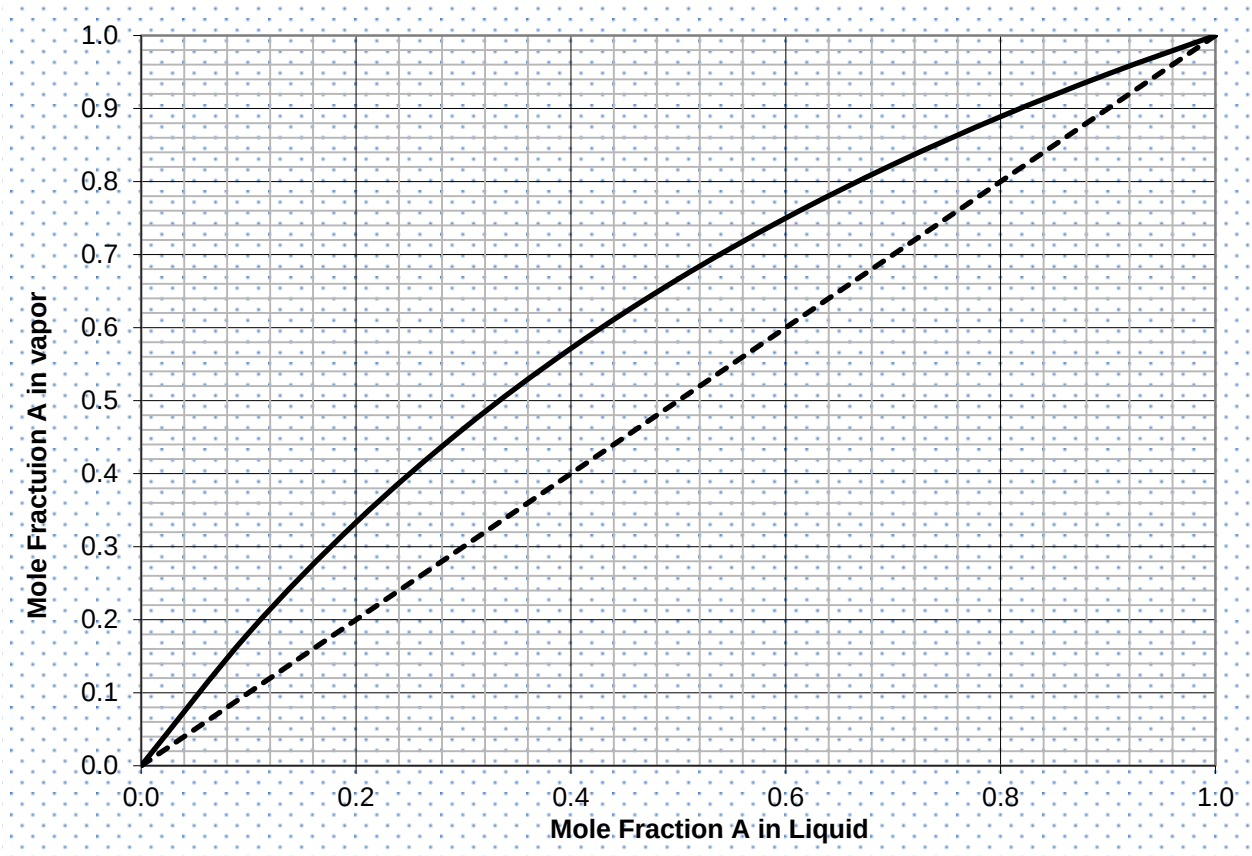
5. ***Distillation Design by McCabe-Thiele Diagram***

A saturated liquid mixture containing 45 mole% component A and the balance B is to be separated by distillation to recover 95% of component A in the distillate with a composition of no more than 3 mole% component B.

a. Draw a schematic of the column showing all the problem data and determine the number of required equilibrium stages and the feed tray location at a reasonable reflux ratio.
b. Determine the required number of equilibrium stages if the feed is saturated vapor.

Compare the number of stages and the flows in the rectifying and stripping sections for cases a and b and speculate on the effect on the cost of the column. The equilibrium line is given in the attached x-y diagram.

Equilibrium data for Problem 5



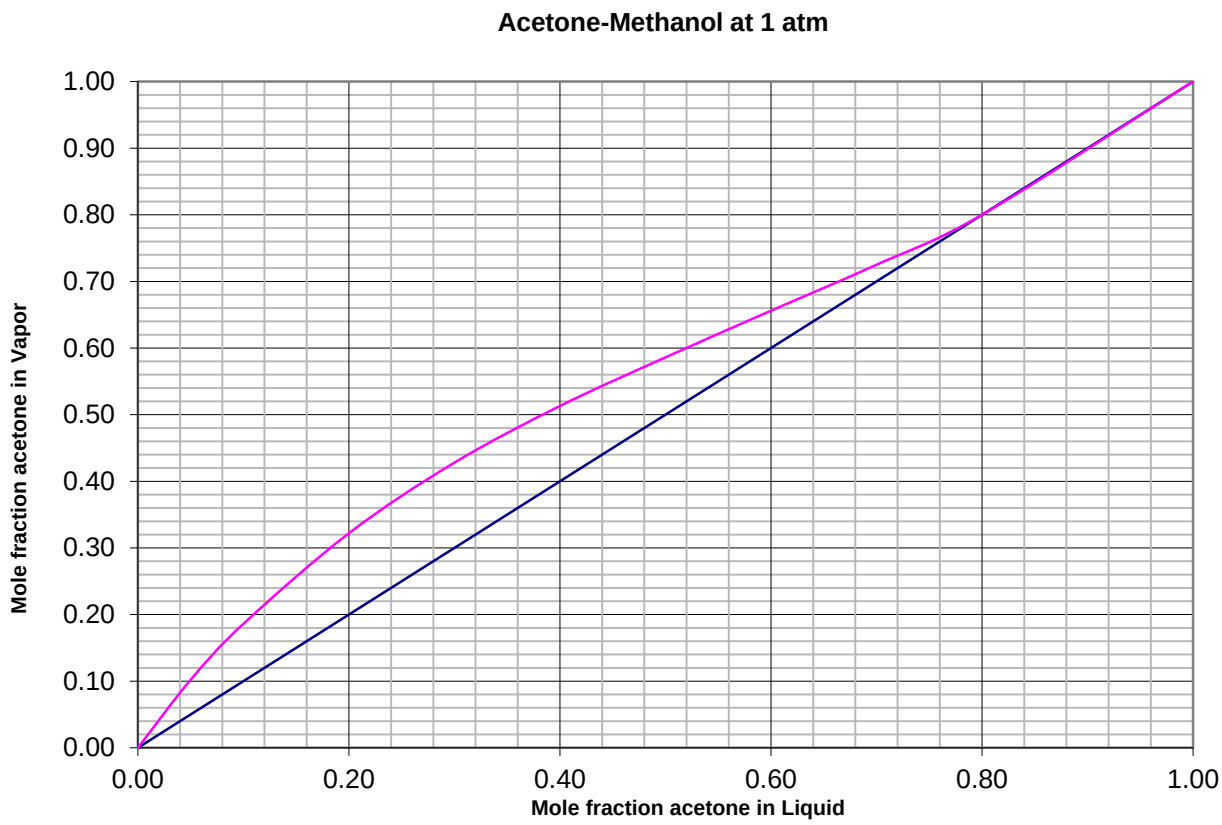
6. ***Design of an Ethanol-Water Column.***

A saturated liquid solution containing 48 weight% ethanol and the balance water is to be separated to recover 99.95% of the ethanol and produce a distillate containing no more than 6 weight% water (190 proof). The feed flow is 1,270 kg/min and the column operates at atmospheric pressure; equilibrium data at that pressure is given in Fig. 20 of the notes. Draw a schematic of the column showing all the problem data and determine the required number of equilibrium stages and the location of the feed stage. Use a reasonable reflux ratio. Report also the liquid and vapor flows in the rectifying and stripping sections of the column.

7. ***Size Ethanol-Water Column.*** For the ethanol-water column of Problem 2, draw a schematic of the column showing all the problem data and determine the condenser heat duty, the reboiler heat duty, the cooling water and steam required in kg/hr, and the column diameter and height in m. Use the results from the McCabe-Thiele solution of Problem 6, and the design premises of Examples 4, 5, 6 and 7.

8. ***Design of a Reboiled Stripping Acetone-Methanol Column.*** A saturated liquid mixture containing 45 weight% acetone and the balance methanol is to be stripped of the acetone by feeding it to the top of a column operating at atmospheric pressure. The stripping column has a reboiler at the bottom to generate the vapors; the distillate product leaves at the top as a vapor and is condensed for further processing, but no reflux is returned to the column. It is desired to have a bottom product containing no more than 2 weight% acetone. Draw a schematic of the column showing all the problem data and determine
   a. the minimum vapor-to-feed ratio,
   b. the required number of equilibrium stages at a vapor rate of 1.2 times the minimum, and
   c. the recovery of the methanol in the bottoms product.

   Report also the product rates and compositions, and the vapor and liquid rates in the column. Equilibrium data for acetone-methanol at atmospheric pressure is given in McCabe, Smith, & Harriott, 7th ed., Table 21.6, page 731. The molecular weights are 58 for acetone, 32 for methanol.



Equilibrium data for Problem 8.

9. **Design of a Refluxed Absorption Methanol-Water Column.** A saturated vapor mixture containing 30 weight% methanol (MW= 32) and the balance water (MW = 18) is fed at a rate of 13,500 kg/hr at the bottom of a column to absorb the water. The absorption column, operating at atmospheric pressure, has a condenser at the top to condense the vapors and return part of the condensed liquid as a reflux to the column, while the balance is the distillate product. The bottom product leaves at the bottom as liquid, and there is no reboiler. It is desired to have a distillate product containing no more than 3 weight% water. Draw a schematic of the column showing the problem data and, using determine
   a. the minimum reflux ratio,
   b. the required number of equilibrium stages at a reasonable reflux ratio, and

c. the recovery of the methanol in the distillate product. Report also the product rates and compositions, and the liquid and vapor rates in the column.

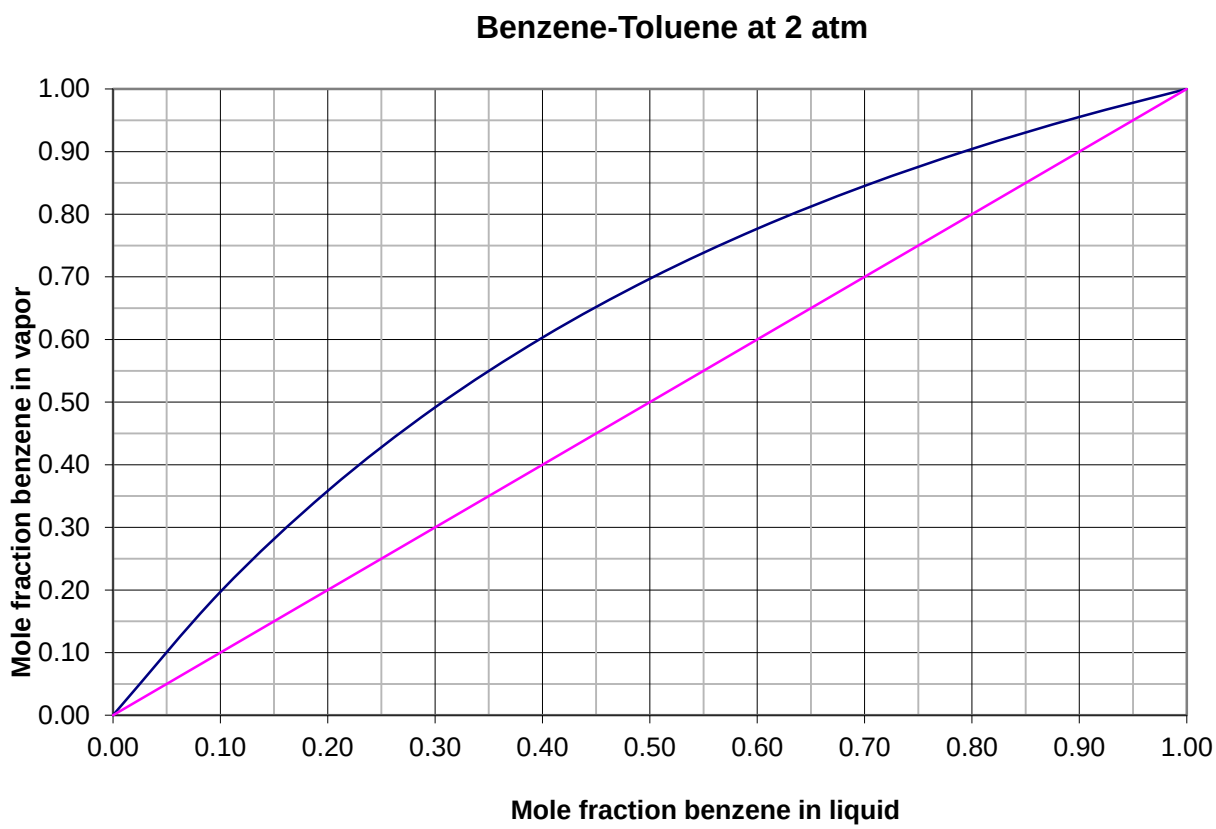
Equilibrium data are given in Fig. 19 of the notes.

10. ***Sizing Methanol-Water Refluxed Absorber***. For the methanol-water column of Problem 9, determine the column height and diameter, the condenser and reboiler duties and areas, the cooling water required in liters/min, and the required steam pressure and flow in kg/hr. Use the design premises of Examples 4, 5, 6, and 7.

11. ***Design of Acetone-Methanol Distillation Column***. A distillation column operating at atmospheric pressure is fed 5,860 kg/min of a saturated vapor mixture containing 30 weight% acetone and the balance methanol. It is desired to recover 90% of the acetone in the distillate product containing no more than 18 weight% methanol. Draw a schematic of the column showing the problem data and determine the required number of actual trays and the optimum location of the feed plate at a reasonable reflux ratio. Assume an over-all column efficiency of 50%, a total condenser and a partial reboiler. Report also the product flows and compositions and the liquid and vapor rates in the rectifying and stripping sections of the column.

The molecular weights are 58 for acetone and 32 for methanol. Equilibrium data are the same as for Problem 8.

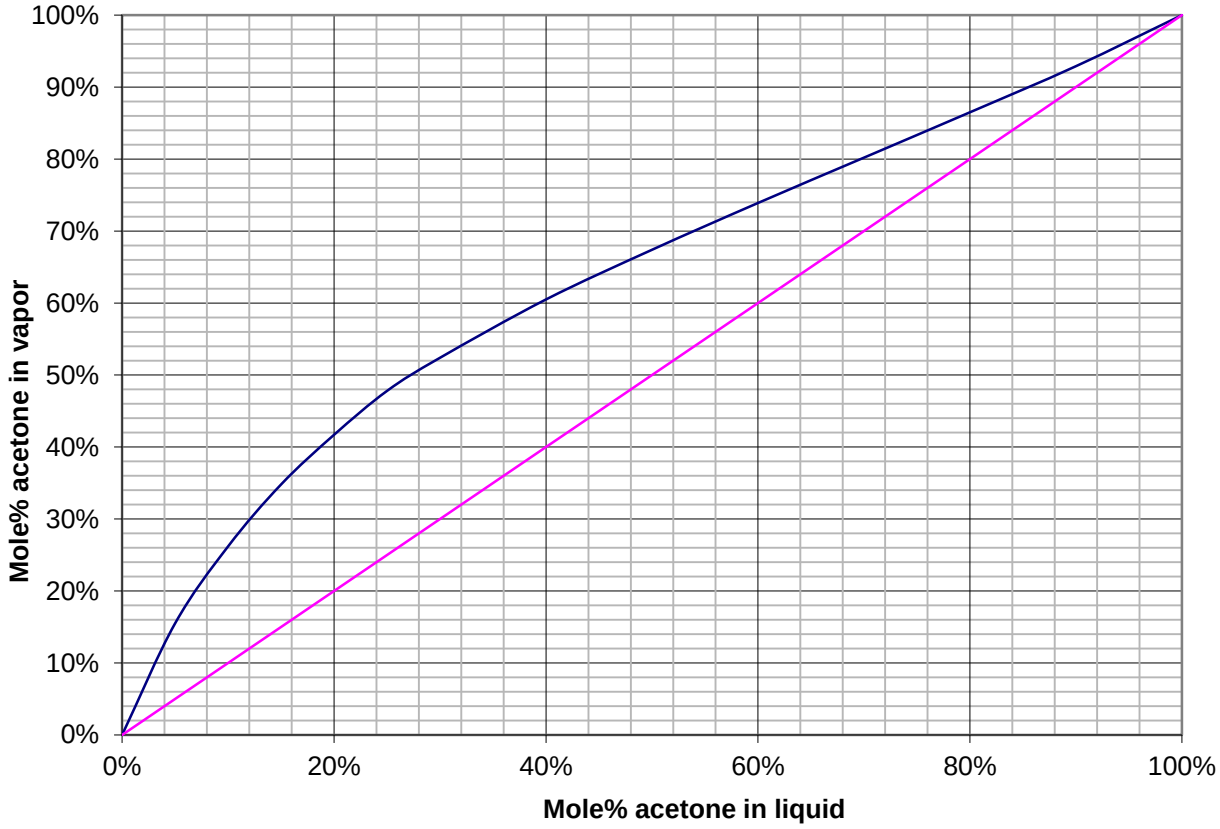
12. ***Design of a Benzene-Toluene Column.*** The feed to a distillation column is 3,630 kmole/hr of a saturated liquid containing 45 mole% benzene and the balance toluene. It is desired to recover 95% of the benzene in the distillate with a composition of no more than 2 mole% toluene. The column is to be run at 2 atm pressure. Draw the schematic of the column showing all the problem data and determine the number of equilibrium stages at total reflux, the minimum reflux ratio, and the required number of equilibrium stages and feed tray location at a reasonable reflux ratio. Report also the product flows and compositions and the liquid and vapor rates in the rectifying and equilibrium sections of the column. The equilibrium diagram at 2 atm is given in the diagram below.



Equilibrium data for Problem 12

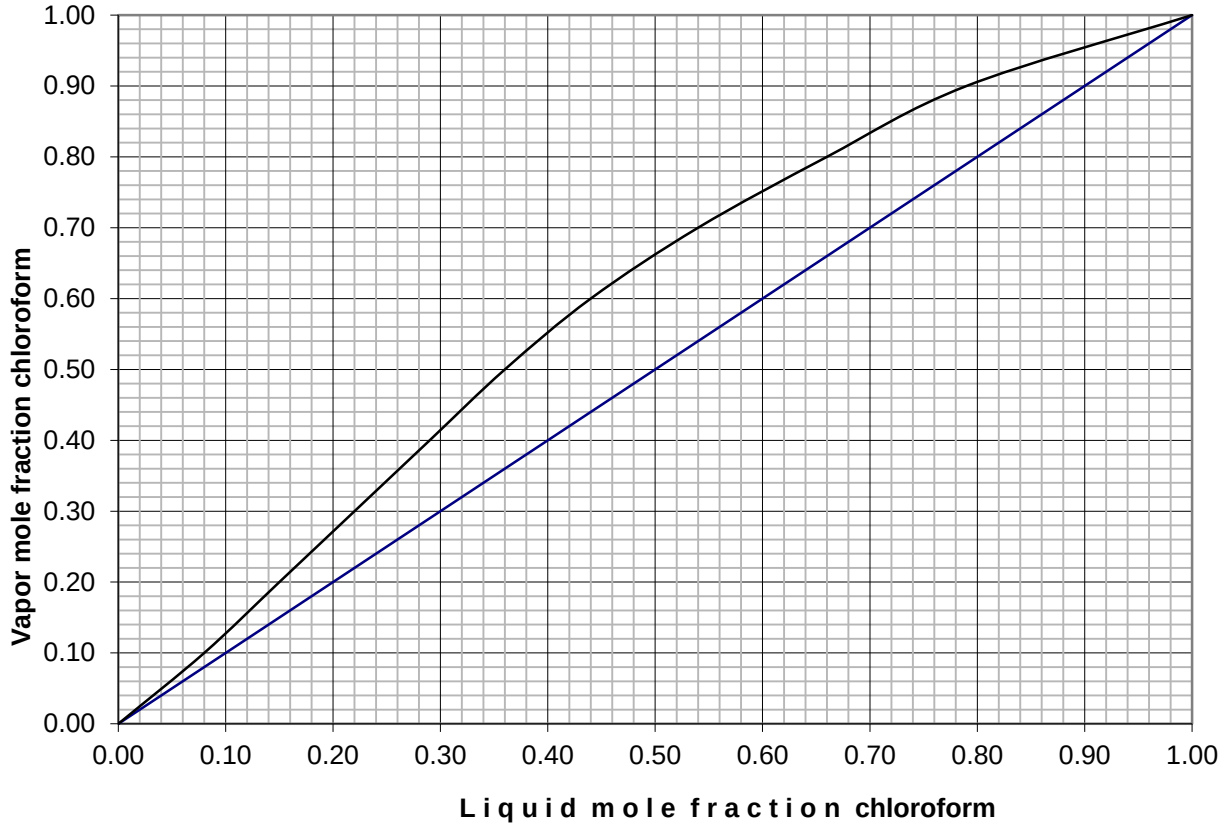
13. ***Sizing of the Benzene-Toluene Distillation Column.*** Size the column of Problem 12. Determine the column height and diameter, the condenser and reboiler duties and areas, the required cooling water rate in liters/min and required steam pressure and flow in kg/hr. Molecular weights are 78 for benzene and 92 for toluene. Use the design premises of Examples 4, 5, 6, and 7.

14. ***Distillation of Acetone-Ethanol Mixture.*** A saturated liquid mixture at a rate of 4,500 kg/hr and containing 45 weight% acetone and the balance ethanol is to be separated by distillation to recover 95% of the acetone in the distillate with a composition of no more than 3 weight% ethanol. The column operates at atmospheric pressure and the x-y diagram at that pressure is attached (obtained from Perry's 7th edition). Draw a schematic of the column showing all the problem data and determine the number of required equilibrium stages and the feed tray location at a reasonable reflux ratio, the product flows and compositions and the liquid and vapor flows in the rectifying and stripping sections of the column.



Equilibrium data for Problem 14

15. ***Design of a Chloroform-Benzene Column.*** A mixture of chloroform and benzene is fed to a distillation column at the rate of 1,600 kmole/hr containing 62 mole% chloroform and the balance benzene. The feed stream is 20% vaporized by mole at the conditions of the column. It is desired to recover 95% of the chloroform in the distillate with a composition of no more than 3 mole% benzene. Equilibrium data at the column pressure of 1 atm is given below. Draw a schematic diagram of the column showing all the problem data and determine the required number of equilibrium stages and feed tray location at a reasonable reflux ratio. Report also the product flows and compositions and the liquid and vapor rates in the rectifying and stripping sections of the column.



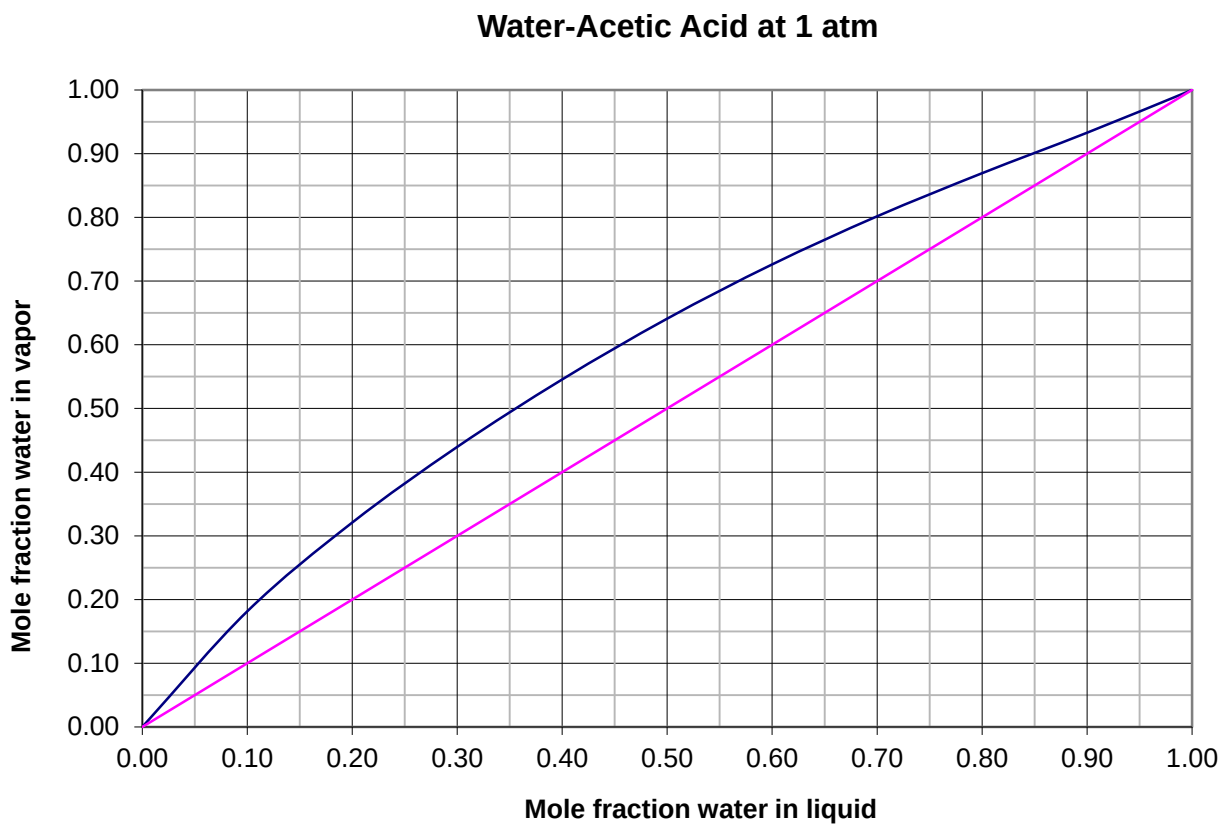
Equilibrium data for Problem 15 (Perry's 3rd ed., Table 1, Page 574)

16. ***Design of a Chloroform-Benzene Refluxed Absorber***. The feed to a refluxed absorber consists of 240 m^3/min at STP (1 atm and 0°C) of a saturated vapor containing 20 mole% chloroform and the balance benzene. It is desired to absorb the benzene out to obtain an overhead product with no more than 2 weight% benzene. The absorber is fed at the bottom and has no reboiler, but the reflux is generated by condensing the overhead vapors and returning part of the condensate. The equilibrium diagram at the column operating pressure of 1 atm is the same as for Problem 15. Draw a schematic of the absorber showing all the problem data and determine the required number of equilibrium stages at a reasonable reflux ratio. Report also the recovery of

chloroform, the product rates and compositions, and the liquid and vapor rates in the column.

17. ***Sizing of the Chloroform-Benzene Distillation Column.*** Size the column of Problem 15. Determine the required number of actual trays, the column height and diameter, the condenser and reboiler heat duties and areas, the cooling water requirement in liters/min, and the required steam, pressure and flow in kg/hr. Use the design premises of Examples 4, 5, 6, and 7.

18. ***Design of a Water-Acetic Acid Column.*** A saturated liquid feed consisting of 3,800 kg/hr containing 75 weight% acetic acid and the balance water is to be separated in a distillation column operating at atmospheric pressure. It is desired to recover 90% of the acetic acid with a purity of no more than 2.5 weight% water. The molecular weights are 60 for acetic acid and 18 for water. Draw a schematic of the column and determine the required number of equilibrium stages and the feed tray location at a reasonable reflux ratio. Report also the product flows and compositions and the liquid and vapor rates in the rectifying and stripping sections of the column. The equilibrium data at the column operating pressure of 1 atm is given below.



Equilibrium data for Problem 18 (Ref: Perry's, 7th ed., Table 13-1, page 13-14.)

19. ***Sizing of Water-Acetic Acid Column.*** Size the column of Problem 18. Determine the required number of actual trays, the column height and diameter, the condenser and reboiler heat duties and areas, the cooling water requirement in liters/min, and the required steam, pressure and flow in kg/hr. Use the design premises of Examples 4, 5, 6, and 7.