Chapter 21. Performance Curves for Individual Unit Operations

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What You Will Learn

- The relationship between input and output when equipment is fixed (already exists) for
  - Fluid flow/pumps/compressors
  - Heat exchangers
  - Tray columns
  - Packed columns
- How to evaluate the performance of the above equipment
- How to represent the performance of the above equipment graphically
- Equipment limitations

As pointed out in the introduction to Section 4, the way in which a process operates will vary significantly throughout its lifetime. Plant operations do not correspond to the conditions specified in the design. This is *not* necessarily a reflection of a poor design. It is a consequence of changes in the process during the life of the plant. There are numerous reasons why a process might not be operated at design conditions. As stated previously, some examples are the following:

- **Design/Construction:** Installed equipment is often oversized. This reduces risks resulting from inaccuracies in design correlations, uncertainties in material properties, and so on.
- **External Effects:** Feed materials, product specifications and flowrates, environmental regulations, and costs of raw materials and utilities all are likely to change during the life of the process.
- **Replacement of Equipment:** New and improved equipment (or catalysts) may replace existing units in the plant.
- **Changes in Equipment Performance:** In general, equipment effectiveness degrades with age. For example, heat transfer surfaces foul, packed towers develop channels, catalysts lose activity, and bearings on pumps and compressors become worn. Plants are shut down periodically for maintenance to restore equipment performance.

With these factors in mind, a good design is one in which operating conditions and equipment performance can be changed throughout the life of the process and plant. This is known as process flexibility. For a company or operation to remain competitive in the marketplace, it must respond to these changes. Therefore, it is essential to understand how equipment performs over its entire operating range and to be able to evaluate the effects of changing process conditions on the overall process performance.

Several techniques for evaluating operating systems are presented in Chapters 19 and 20. The purpose of this chapter is to apply these techniques to obtain a solution to a performance problem for a specific piece of equipment. This solution may be a single answer. However, a much more useful solution is a curve or a family of curves that represents the way an existing piece of equipment or system responds to changes in input or equipment variables. These curves are referred to as **performance curves**. They are the basis for predicting the behavior of existing equipment. Performance curves present a range of possible solutions rather than merely a single answer. In principle, performance equations could be used instead of performance curves. However, by representing equipment performance in graphical form, the performance characteristics are easier to visualize, providing a better intuitive understanding. For example, in Figure 20.1, the boundary for fully developed turbulent flow, above which the friction factor is constant, is more effectively represented by the dotted curve on the graph than by the equation of that curve.

Performance curves represent the relationship between process outputs and process inputs.

By plotting the response variables as a function of the input variables, the sensitivity of one to the other becomes immediately obvious. Such sensitivity cannot be inferred from a single (numerical) solution. With the wide availability of spreadsheets and process simulators, the effort expended generating performance curves may be justified.

In order to construct performance and system curves, material and energy balance equations must be used along with the (design) equations relating equipment parameters. However, other constraints may also have to be considered, such as the maximum or minimum system temperature and pressure allowed for the equipment, the maximum velocity of fluid through the equipment to avoid excessive erosion, the maximum or minimum velocity to avoid flooding in packed towers and tray columns, the minimum velocity through a reactor to avoid defluidization, the maximum residence time in a reactor to avoid coking/cracking reactions or by-product formation, the minimum flow through a compressor to avoid surging, or the minimum approach temperature to avoid the condensation of acidic gases inside heat exchangers.

It is possible to construct performance curves for essentially any piece of equipment. In this chapter, performance curves are developed for several specific equipment types. This case-study approach is used to help you develop your own methods for generating performance and system curves for equipment not covered in this text.

21.1. Application to Heat Transfer

In this section, the steam generator shown in Figure 21.1(a) is analyzed to illustrate the preparation and value of a performance diagram for predicting the response of a heat exchanger to changing conditions. The data shown were obtained from an actual operating system. A steam generator is similar to a shell-and-tube heat exchanger. Figure 21.2 contains drawings of portions of shell-and-tube heat exchangers.

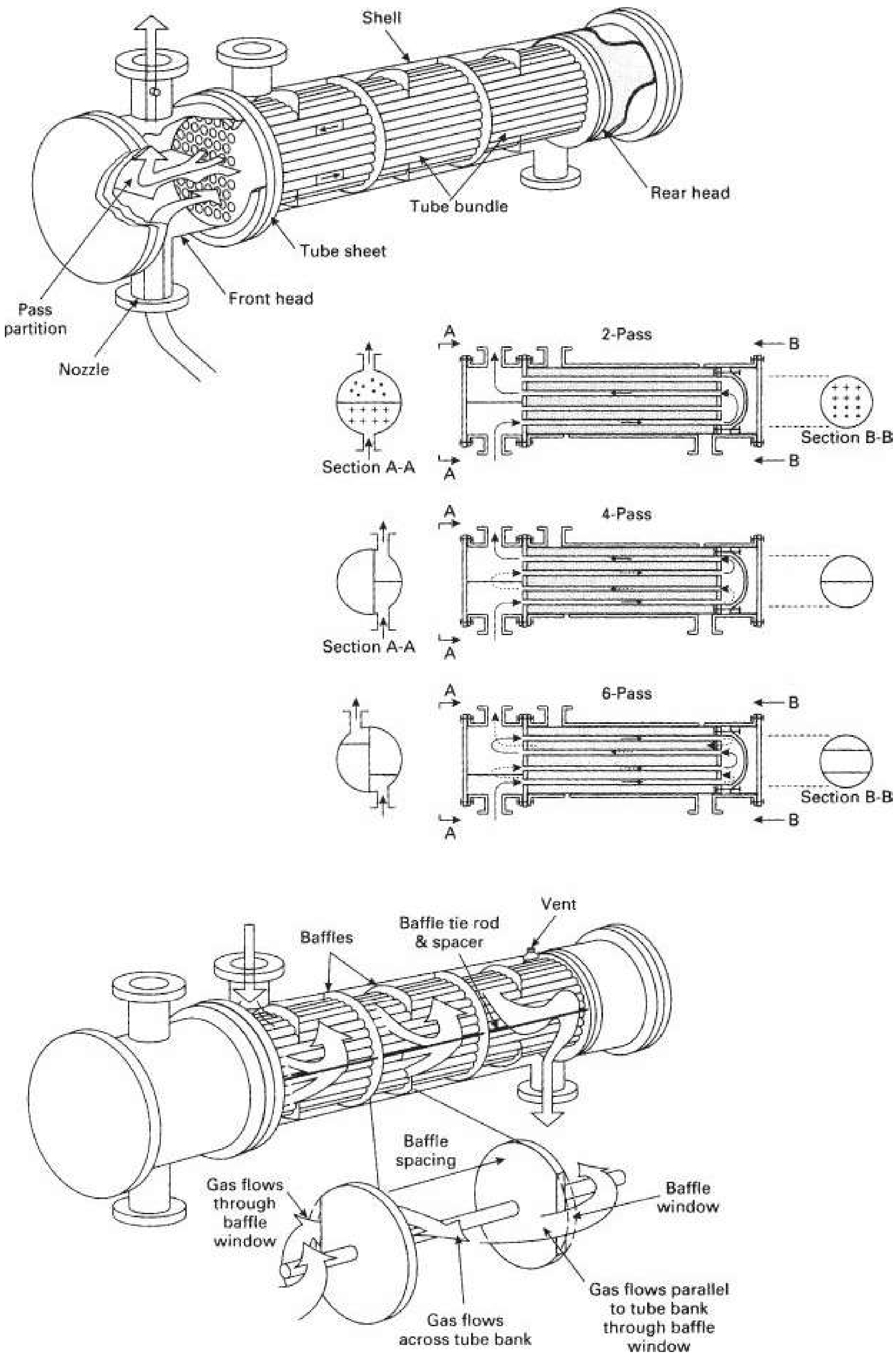

Figure 21.2. Details of a Shell-and-Tube Heat Exchanger (From D. R. Woods, *Process Design and Engineering Practice*, Englewood Cliffs, NJ: Prentice Hall, 1995)

Saturated steam is produced in a kettle-type vaporizer containing long vertical tubes. Heat is provided from a hot light oil stream that enters at 325°C and leaves at 300°C. The effective area for heat transfer is adjusted by changing the level of the boiling liquid in the exchanger.

Figure 21.1a provides the current operating conditions, some limited thermodynamic data, data on the vaporizer, and a sketch of the equipment.

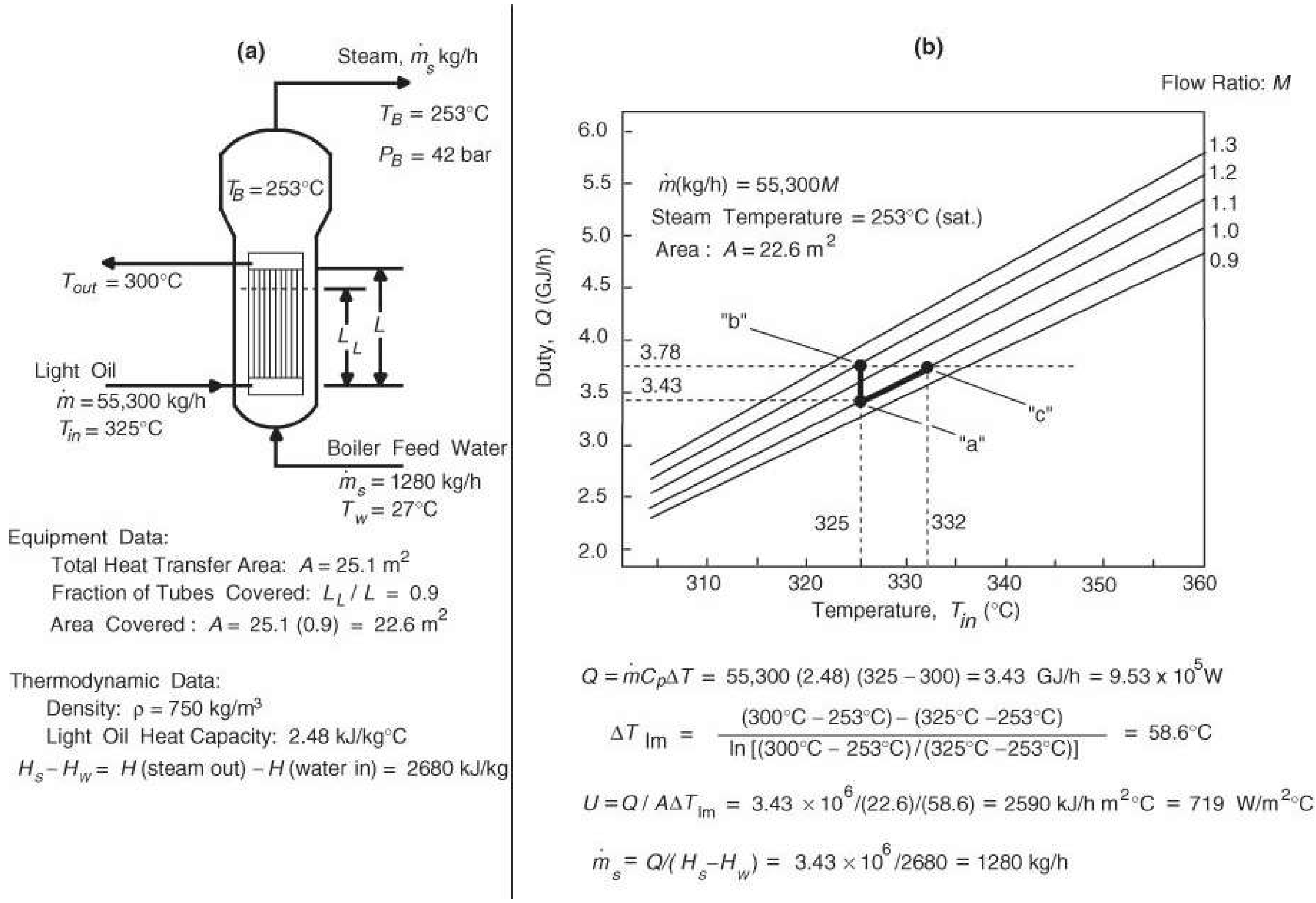

Figure 21.1. Performance Diagram for a Heat-Exchange System

The conditions given in Figure 21.1(a) designate the base case. Development of performance curves involves solving the following three equations simultaneously:

$$\text{Heat lost by light oil, } Q = \dot{m}(H_1 - H_2) = \dot{m}C_{p,\,oil}(T_{in} - T_{out}) \tag{21.1}$$

$$\text{Heat gained by water stream, } Q = \dot{m}_s(H_s - H_w) = \dot{m}_s[C_p(T_B - T_w) + \lambda_s] \tag{21.2}$$

$$\text{Heat transferred, } Q = UA\Delta T_{lm} \tag{21.3}$$

In these equations,

H_1 and H_2 refer to the specific enthalpy, kJ/kg, for the light oil, with the subscripts 1 for inlet and 2 for outlet conditions.

H_s and H_w refer to the specific enthalpy, kJ/kg for the steam and water, respectively.

$\dot{m}$ and $\dot{m}_s$ refer to the stream flowrate, kg/h, for the light oil and water/steam stream, respectively.

λ_s is the latent heat of the steam.

The heat transfer equation, Equation (21.3), contains the characteristic factors specific to the heat transfer equipment.

Figure 21.1(b) shows the calculations used for the base case. They include the heat transfer rate, Q, the steam generation rate, $\dot{m}_s$, the overall heat transfer coefficient, U, and the log-mean temperature difference, ΔT_{lm}.

Figure 21.1(b) is the performance diagram for this boiler system subject to the following constraints:

1. The temperature of the boiling water, T_B = 253°C, which is set by the pressure of the water in contact with the tubes
2. The liquid level, which is 0.9L (where L = tube height)
3. Total heat transfer area, A_T = 25.1 m^2
4. The time at which the operating data were obtained

The heat transfer rate or vaporization duty, Q, is plotted in Figure 21.1(b) as a function of the inlet light oil coolant temperature, T_{in}, with flowrate ratio of light oil, M, as a parameter. M is the ratio of the current oil flowrate (subscript 2) to the base-case oil flowrate (subscript 1).

$$M = \dot{m}_2/\dot{m}_1 \qquad (21.4)$$

where $\dot{m}_1 = 55{,}300\ \text{kg/h}$.

Before discussing how these curves were developed, Example 21.1 illustrates how the performance curves can be used to predict the effects of changing operating conditions.

Example 21.1.

An increase in steam production of 10% is needed. You are to provide the operator with the new input stream conditions for two cases by completing the following table:

	Light Oil Flow (Mg/h)	T_{in}(°C)	T_{out} (°C)
Current Values	55.3	325	300
Case (a)	55.3	?	?
Case (b)	?	325	?

These problems can be solved by the methods developed in Chapter 20. If only a single answer is desired, this method is faster than preparation of a complete performance curve, which is described later.

If the steam production must increase by 10%, then the ratio of the new case (2) to the base case (1) from Equation (21.2) is

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_{s2}(H_s - H_w)_2}{\dot{m}_{s1}(H_s - H_w)_1} = 1.1 \qquad \text{(E21.1a)}$$

Because the enthalpy difference is unchanged for the phase change of water to steam, the ratio Q_2/Q_1 = 1.1. Now, the ratio of Q_2/Q_1 must be written for the remaining two equations, (21.1) and (21.3). For Case (a), these are

$$\frac{Q_2}{Q_1} = 1.1 = \frac{\dot{m}_2 C_{p2}(T_{in} - T_{out})_2}{\dot{m}_1 C_{p1}(T_{in} - T_{out})_1} = \frac{(T_{in} - T_{out})_2}{25} \qquad \text{(E21.1b)}$$

$$\frac{Q_2}{Q_1} = 1.1 = \frac{U_2 A_2 \Delta T_{lm2}}{U_1 A_1 \Delta T_{lm1}} = \frac{(T_{out} - T_{in})_2}{58.6 \ln\left(\dfrac{T_{out2} - 253}{T_{in2} - 253}\right)} \frac{U_2}{U_1} \qquad \text{(E21.1c)}$$

Because the mass flowrate of the oil does not change, its heat transfer coefficient remains constant. Because a boiling heat transfer coefficient does not change with flowrate, the overall heat transfer coefficient is unchanged. Therefore, $U_1 = U_2$. These assumptions were used in obtaining Equation (E21.1[c]). Solving Equations (E21.1[b]) and (E21.1[c]) simultaneously yields

$$T_{in2} = 332°\text{C}$$

$$T_{out2} = 304.5°C$$

For Case (b), it is necessary to know more detail about the heat transfer coefficient, because the mass flowrate of the oil changes. For this system, it is known that the boiling heat transfer coefficient (h_o) is two times the oil heat transfer coefficient (h_i). Because only the oil heat transfer coefficient changes with flowrate, assuming negligible fouling and wall resistances, the overall heat transfer coefficient is expressed in terms of the oil heat transfer coefficient as

$$\frac{1}{U_1} = \frac{1}{h_{i1}} + \frac{1}{h_{o1}} = \frac{3}{2h_{i1}} \quad \text{(E21.1d)}$$

$$\frac{1}{U_2} = \frac{1}{h_{i1}M^{0.8}} + \frac{1}{h_{o2}} = \frac{1}{h_{i1}}\left(\frac{1}{M^{0.8}} + 0.5\right) \quad \text{(E21.1e)}$$

where $M = \dot{m}_2/\dot{m}_1$. The base-case ratios now become

$$1.1 = \frac{M(325 - T_{out2})}{25} \quad \text{(E21.1f)}$$

$$1.1 = \frac{3(T_{out2} - 325)}{2(58.6)\left(\frac{1}{M^{0.8}} + 0.5\right)\ln\left(\frac{T_{out2} - 253}{72}\right)} \quad \text{(E21.1g)}$$

Solving these two equations simultaneously yields T_{out2} = 301.5°C and M = 1.17, which means that the flowrate of oil must be increased by 17% to obtain a 10% increase in steam production, and the resulting outlet oil temperature rises slightly.

The same results can be obtained from the performance graph in Figure 21.1(b).

New Steam Production, $\dot{m}_s$ = 1.1(1280) = 1410 kg/h

New Exchanger Duty, $Q = \dot{m}_s (H_s – H_w)$ = 1410(2680) = 3.78 × 10^6 kJ/h

Case (a): On Figure 21.1(b), the base-case operating condition is point "a." From point "a," follow the constant flow line, M = 1.0, to a vaporizer duty, Q = 3.78 × 10^6 kJ/h. This is line segment "a"–"c," and, from this, the inlet oil temperature, T_{in2} = 332°C.

From Equation (21.1)

$$T_{out2} = T_{in2} - Q/(\dot{m}_2 C_{p,oil}) = 332 - 3.78 \times 10^6/[55{,}300(2.48)] = 305°C$$

Case (b): On Figure 21.1(b) follow the constant temperature line for T_{in2} = 325°C to the new vaporizer duty, Q = 3.78 × 10^6 kJ/h. This is shown as line segment "a"–"b." From this, the value of M = 1.17.

$$\dot{m} = 1.17(55{,}300) \ = \ 64{,}700\ \text{kg/h (17\% increase in flow)}$$

From Equation (21.1)

$$T_{out2} = T_{in2} - Q/(\dot{m}_2 C_{p,oil}) = 332 - 3.78 \times 10^6/[64{,}700(2.48)] = 301.5°C$$

In Example 21.1, two sets of operating conditions were calculated that would provide the required increase in steam production. The input oil flowrate, $\dot{m}$, could be increased with the temperature, T_{in}, held constant, or the oil flowrate, $\dot{m}$, could be held constant while the input oil temperature, T_{in}, is increased. These two solutions, along with an infinite number of other solutions, lie on a line of constant heat duty, Q = 3.78 GJ/h.

The curves for constant flow ($\dot{m}$ = 55,300M) in Figure 21.1(b) are straight lines. Equating and rearranging Equations (21.1) and (21.3) produces

$$(T_{in} - T_B)/(T_{out} - T_B) = K \quad \text{(21.5)}$$

$$\text{where } K = \exp[UA/(\dot{m}/C_{p,oil})] \tag{21.6}$$

By solving Equation (21.5) for T_{out} and substituting it into Equation (21.1),

$$Q = \dot{m}C_{p,oil}(T_{in} - T_B)[1 - (1/K)] \tag{21.7}$$

The overall heat transfer coefficient, U, varies with the light oil flowrate, $\dot{m}$. The performance for a constant value of oil flowrate is obtained once the value of K is known.

Equation (21.7) was used to obtain the curves given in Figure 21.1(b). The small amount of heat transferred above the liquid level (where the heat transfer coefficient for the gas phase will be small) was ignored. In addition, it was assumed that the incoming feed water mixes rapidly with the large volume of boiling liquid water in the vaporizer. Thus, the water outside the tubes is essentially constant and equal to the saturation temperature. The simple form of Equation (21.7) is a consequence of this constant temperature. When neither stream involves a phase change, the evaluation is more complicated. One reason is the need to consider the log-mean temperature correction factor for these types of shell-and-tube heat exchangers.

In reality, fouling will affect heat-exchanger performance over time. **Fouling** is a buildup of material on tube surfaces, and it occurs to some extent in all heat exchangers. When a heat exchanger is started up, there will be no fouling; however, with time, trace impurities in the fluids deposit on the heat exchanger tubes. This is usually more significant for liquids, and among the possible impurities are inorganic salts or microorganisms. The fouling layer provides an additional resistance to heat transfer. For a heat exchanger constructed from material with high thermal conductivity operating with fluids having high heat transfer coefficients, fouling may provide a greater resistance than the convective film resistance. The influence of fouling on heat-exchanger performance is the subject of problems at the end of the chapter.

Which of the two heat transfer surfaces provides the major fouling resistance in the steam generator? The water outside the tubes is designated as boiler feed water. This stream is an expensive source of water. It has been treated extensively to remove trace minerals and hence reduces significantly the fouling of heat transfer surfaces. This suggests that fouling on the outside of the tubes will remain low and the oil stream is the major contributor to fouling.

Example 21.2 suggests ideas for alternative performance of the steam generator. Some of these are the subject of problems at the end of the chapter.

Example 21.2.

Assume that neither the oil stream flowrate, $\dot{m}$, nor the inlet temperature, T_{in}, can be changed. What can be done to increase steam production, $\dot{m}_s$?

The heat transfer equation, $Q = UA\Delta T_{\text{lm}}$, shows that Q can be increased by

- **a. Increasing U:** Clean the tubes.
- **b. Increasing A:** Increase the liquid level in the boiler.
- **c. Increasing ΔT_{lm}:** Decrease the boiler temperature by lowering the pressure.
- d. Combinations of (a), (b), and (c).

21.2. Application to Fluid Flow

In this section, performance curves for centrifugal pumps, reciprocating pumps, and piping networks are discussed. The use of pump and system curves is introduced and the concepts of net positive suction head are discussed.

21.2.1. Pump and System Curves

In this section, performance curves for a centrifugal pump and the flow of a liquid through a pipe network connecting a storage tank to a chemical reactor are presented. Centrifugal pumps are very common in the chemical industry. Figure 21.3 shows the inner workings of a centrifugal pump. It is important to understand that the performance curves presented here are unique to centrifugal pumps. Positive displacement pumps, the other common type of pump used in the chemical industry, have a completely different performance curve and are discussed briefly in Section 21.2.3.

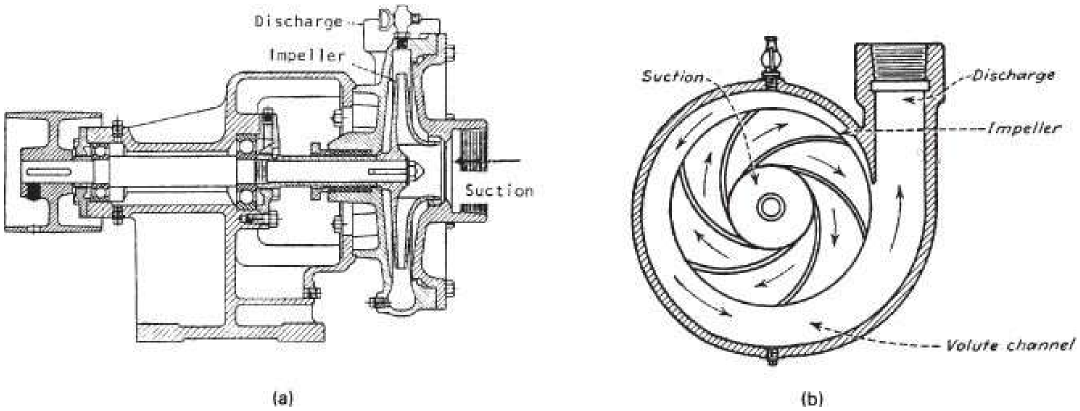

Figure 21.3. Inner Workings of a Centrifugal Pump (From S. Walas, *Chemical Process Equipment: Selection and Design.* Stoneham, MA: Butterworth, 1988. Reproduced with permission.)

In situations involving the flow of fluid, the equation used to relate pressure changes and flowrate is the mechanical energy balance (or the extended Bernoulli equation):

$$\frac{\Delta P}{\rho g} + \frac{\Delta u^2}{2g} + \Delta z = \frac{W_s}{g} - \frac{F_d}{g} \tag{21.8}$$

The terms shown on the left-hand side of Equation (21.8) are point properties. They are independent of the path taken by the fluid. The Δ in Equation (21.8) represents the difference between outlet and inlet conditions on the control volume. In contrast, the terms on the right-hand side of the equation are **path properties** that depend on the path followed by the fluid. Work is defined as positive when done on the system. The terms related to the path taken by the fluid are specific to the operating system and form the basis for the performance curves.

Equation (21.8) is written in terms of fluid head. Each term has units of length. (The equation could be written so that each term has units of pressure by multiplying each term by ρg.) **Fluid head** is a way of expressing pressure as an equivalent static pressure of a stationary body of fluid. For example, one atmosphere of pressure is equivalent to about 34 feet of water, because the pressure difference between the top and bottom of a column of 34 feet of water is one atmosphere. Equation (21.8), with each term defined as fluid head, is

$$\Delta h_p + \Delta h_v + \Delta h_z = h_s - h_f \tag{21.9}$$

A centrifugal pump is shown in Figure 21.4 along with its performance diagram. In the performance diagram, the ordinate is in units of pressure head, which is how all of the curves discussed in this section are usually represented. However, it is also possible to represent these curves in pressure units. In this book, examples and problems using both pressure units and head units are included. For this analysis, the change in the velocity head and elevation head between points “1” and “2” is either zero or small, and Equation (21.9) reduces to

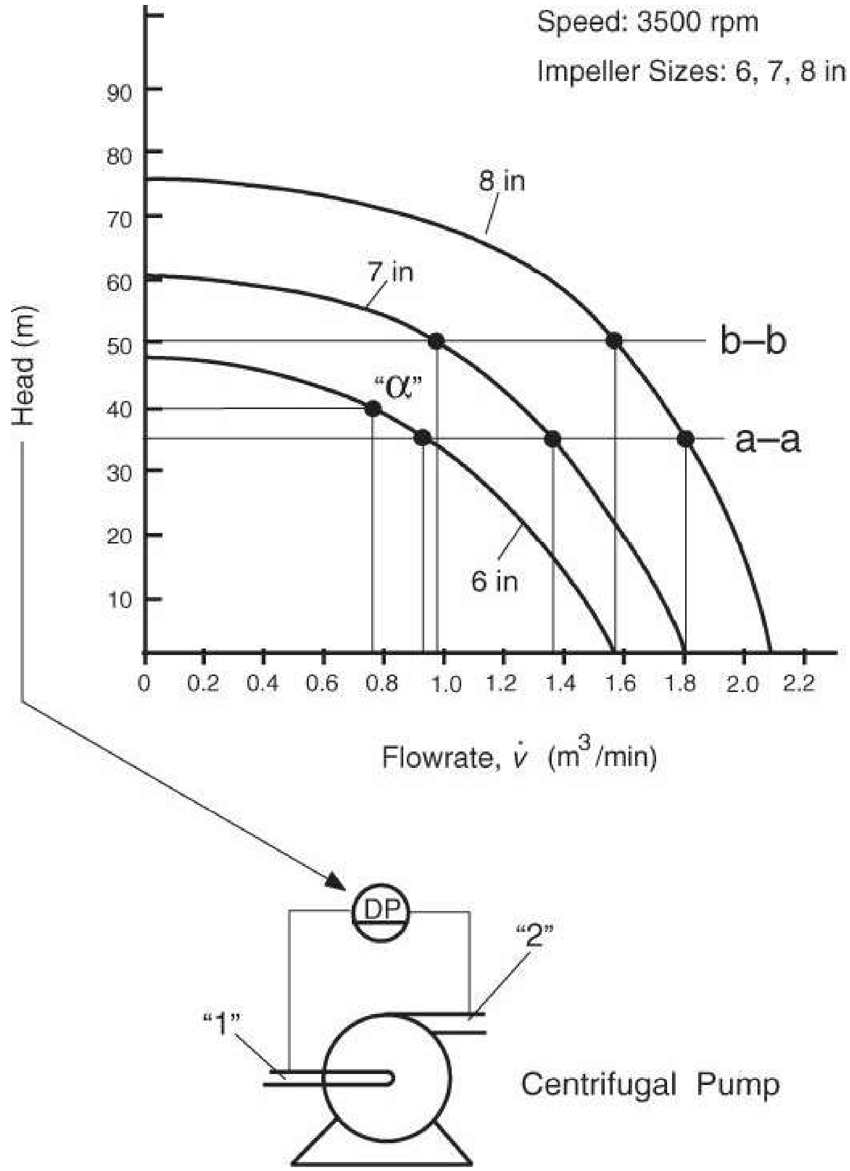

Figure 21.4. Performance Diagram for a Centrifugal Pump

$$\Delta h_p = h_s - h_f \qquad (21.10)$$

Centrifugal pumps are used in a variety of applications and are available from many manufacturers. Pump impellers may have forward or backward vanes, may have enclosed or open configurations, and may operate at different speeds. Their performance characteristics depend on their mechanical design. As a result, experimentally determined performance curves are supplied routinely by the pump manufacturer.

The performance curve, or pump curve, shown in Figure 21.4 is typical of a centrifugal pump handling normal (Newtonian) liquids. It relates the fluid flow-rate to the head, h_s. To understand this relationship, consider point "α" shown on the performance curve in Figure 21.4. This point lies on the curve for a 6-in impeller. It shows that the pump delivers 0.76 m^3/min of liquid when the pressure differential between the suction and discharge side of the pump is equivalent to a column of liquid 40 m in height.

Figure 21.4 contains three curves, each depicting the pump performance with a different impeller diameter. The pump casing can accept any of these three impellers, and each impeller has a unique pump curve. Pump curves often display efficiency and horsepower curves in addition to the curves shown in Figure 21.4.

From the pump curves in Figure 21.4, the following trends can be seen for each impeller:

1. The head produced at a given flowrate varies (increases) with the impeller diameter.
2. The pump provides low flowrates at high head and high flowrates at low head.
3. The head produced is sensitive to flowrate at high flowrates; that is, the curve drops off sharply at high flowrates.
4. The head produced is relatively insensitive to flowrate at low flowrates; that is, the curve is relatively flat at low flowrates.

Example 21.3 demonstrates how to read a pump curve.

Example 21.3.

The centrifugal pump shown in Figure 21.4 is used to supply water to a storage tank. The pump inlet is at atmospheric pressure and water is pumped up to the storage tank, which is open to atmosphere, via large-diameter pipes. Because the pipe diameters are large, the frictional losses in the pipes and any change in fluid velocity can be safely ignored.

a. If the storage tank is located at an elevation of 35 m above the pump, predict the flow using each impeller.
b. If the storage tank is located at an elevation of 50 m above the pump, predict the flow using each impeller.

Solution

a. From Figure 21.4, at Δh_p = 35 m (see line “a–a”)

6-in Impeller: Flow = 0.93 m^3/min

7-in Impeller: Flow = 1.38 m^3/min

8-in Impeller: Flow = 1.81 m^3/min

b. From Figure 21.4, at Δh_p = 50 m (see line “b–b”)

6-in Impeller: Flow = 0 m^3/min

7-in Impeller: Flow = 0.99 m^3/min

8-in Impeller: Flow = 1.58 m^3/min

In Example 21.3, the flowrate into the tank is restricted to one of three discrete values (depending on the impeller installed in the pump housing). For the system described, once the elevation of the tank and the impeller diameter are chosen, a single unique flowrate is obtained. Other flowrates cannot be obtained for this system.

Example 21.3 demonstrates the need to know the characteristic shape of the performance curve at the current operating point to predict the effect of any change in tank elevation. In Example 21.3, because the 8-in impeller is operating in a region where the flow is not greatly affected by an increase in head, the flowrate was reduced by less than 15%. In contrast, for the 6-in impeller, where a change in head has a large effect, the flowrate is reduced by 100% (no fluid flows).

It is essential to understand the system performance before making predictions or recommendations.

Consider the flow situation shown in Figure 21.5. The figure illustrates a pipe network through which feed material is transported from a storage tank through a heat exchanger to a chemical reactor. From the information provided on Figure 21.5, it is possible to construct the performance curve shown, which is called the **system curve**.

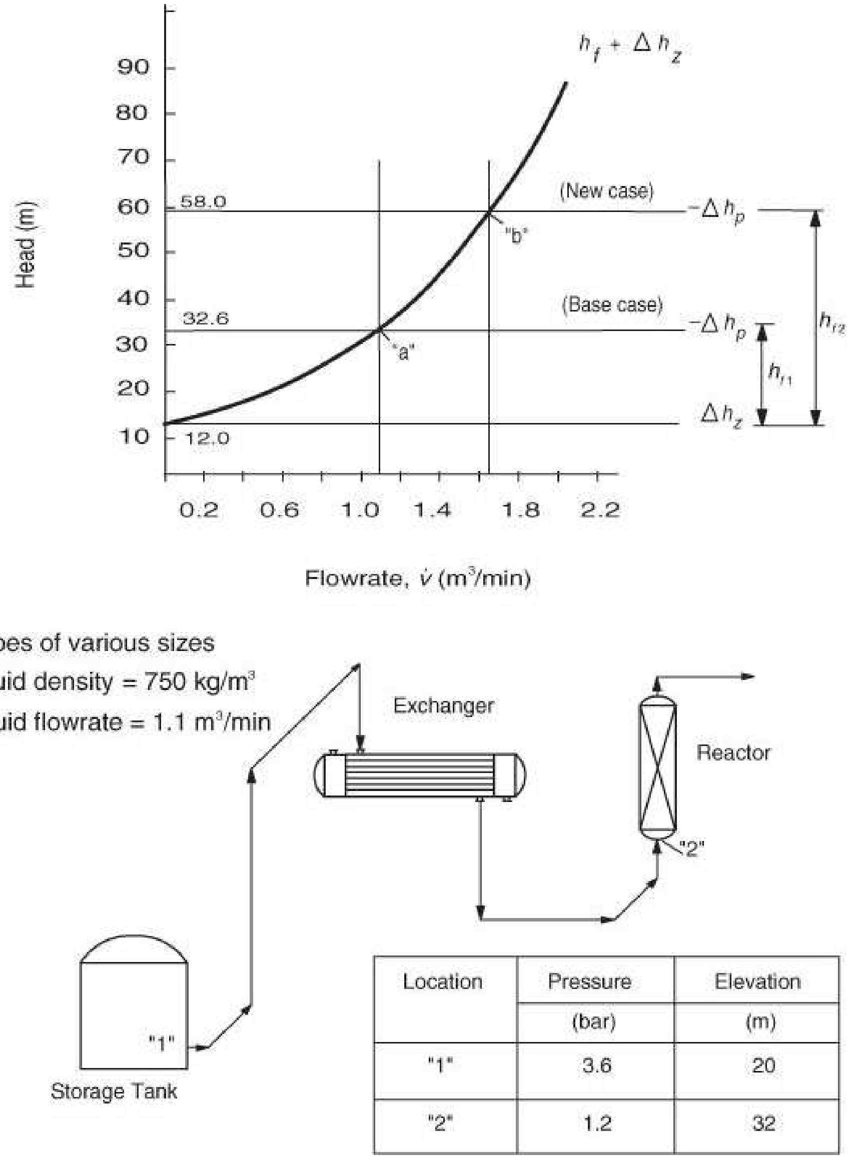

Figure 21.5. Sketch and Performance Diagram for an Operating Flow System

Because the fluid is a liquid, it can be assumed that the change in velocity head, Δh_v, is small. In addition, the work head $\Delta h_s = 0$, and Equation (21.9) written between points "1" and "2" reduces to

$$\Delta h_p = -h_f - \Delta h_z \qquad (21.11)$$

The pressure head term, Δh_p, and the elevation head term, $-\Delta h_z$, are constant and independent of the flowrate. The friction term, $-h_f$, in Equation (21.11) depends upon

1. The specific system configuration

2. The flowrate of fluid

For this system, for fully developed turbulent flow, the relationship needed to predict the friction term, h_f, as a function of fluid velocity is known. The relationship is given in Table 20.1 and in Example 20.2. This relationship in terms of fluid head is

$$h_{f2} = h_{f1}(u_2/u_1)^2 = h_{f1}(\dot{v}_2/\dot{v}_1)^2 \qquad (21.12)$$

where $\dot{v}$ is the volumetric flowrate, the subscript 2 refers to the new case, and the subscript 1 refers to the base case. Equation (21.12) represents a parabola, which is the shape of the system curve in Figure 21.5. In terms of pressure, Equation (21.12) becomes

$$\Delta P_2 = \Delta P_1(\dot{v}_2/\dot{v}_1)^2 \qquad (21.13)$$

and it is also possible to plot the system curve in terms of pressure, as is done later in Figure 21.6.

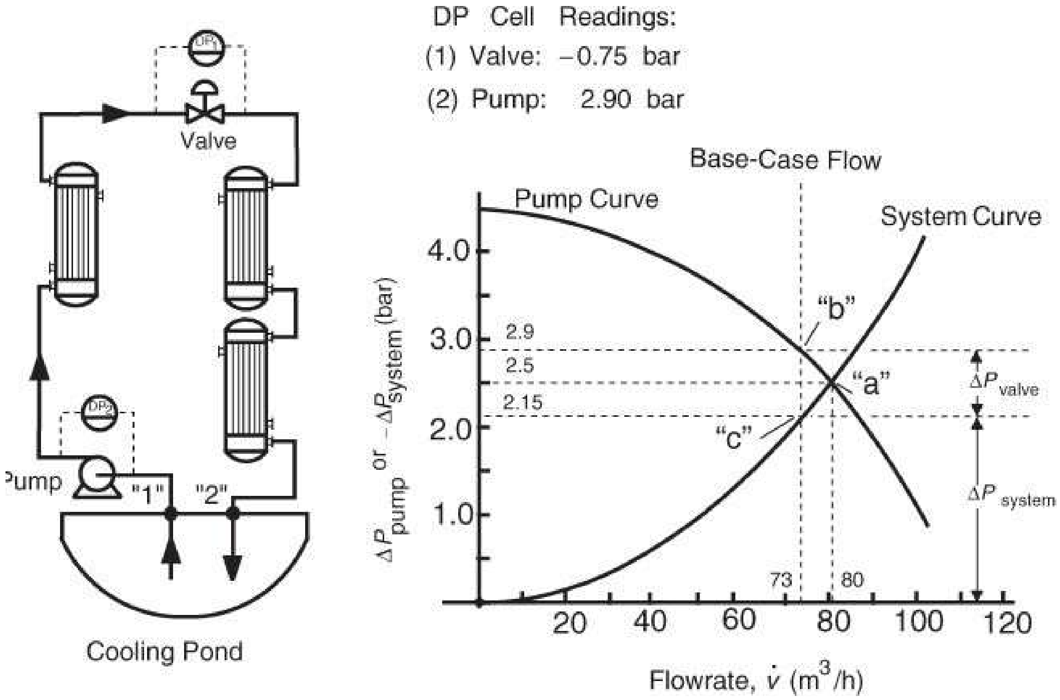

Figure 21.6. Performance Curves for Coolant System

The operating conditions given in Figure 21.5 serve as the base case in Examples 21.4 and 21.5, which illusrate the ratio method discussed previously.

Example 21.4.

The fluid flowrate measured for the conditions shown in Figure 21.5 is 1.1 m^3/min.

a. Using the information given on the flow diagram (see Figure 21.5) for the base case, determine the value of h_f.

b. Develop the equation for h_f as a function of the flowrate, $\dot{v}$.

Solution

a. From Equation 21.11,

$$h_f = \Delta - h_z - \Delta h_p$$

$\Delta h_z = z_2 - z_1 = 32$ m – 20 m = 12 m (Δh_z is drawn as a horizontal line in Figure 21.5)

$$\Delta h_p = (P_2 - P_1)/\rho g = (1.2 - 3.6) \times 10^5/[750(9.81)] = -32.6\text{ m}$$

$$h_f = -12\text{ m} + 32.6\text{ m} = 20.6\text{ m}$$

b. $h_{f2} = h_{f1}(\dot{v}_2/\dot{v}_1)^2 = (20.6\text{ m})(\dot{v}_2/1.1)^2 = 17.02\dot{v}_2^2$

Example 21.5.

The flow to the reactor in Example 21.4 is increased by 50%. The pressure of the reactor is held constant. Determine the pressure required at the exit of the storage tank.

For a 50% increase, $\dot{v}_2 = 1.5\,\dot{v}_1 = 1.1(1.5) = 1.65\text{ m}^3\text{/min}$.

The line of constant flow, $\dot{v} = 1.65\text{ m}^3\text{/min}$, is shown as the vertical line through point "b" in Figure 21.5. It intersects the performance line, $h_f + \Delta h_z = 58$ m (point "b").

$$-\Delta h_p = 58\text{ m} = (P_1 - P_2)/\rho g = P_1/\rho g - 1.2 \times 10^5/[750(9.81)] = P_1/\rho g - 16.3\text{ m}$$

$$P_1 = (58 + 16.3)\rho g = 74.3(750)(9.81) = 5 \times 10^5\text{ Pa} = 5.47\text{ bar}$$

In this section, two representative types of performance curves were considered. For the pump, the information needed for developing a performance diagram was obtained experimentally and supplied by the manufacturer. In the flow network, a base case was established using actual operating data and the performance diagram, known as the system curve, calculated using the mechanical energy balance.

21.2.2. Regulating Flowrates

In all systems presented in this chapter, input flowrates are the primary variables that are used to change the performance of a system. In fact, in a chemical plant, process regulation is achieved most often by manipulating flowrates, which is accomplished by altering valve settings. If it is necessary to change a temperature, the flow of a heating or cooling medium is adjusted. If it is necessary to change a reflux ratio, a valve is adjusted. In this section how to regulate these input flows to give desired values is considered.

Process conditions are usually regulated or modified by adjusting valve settings in the plant.

Although valves are relatively simple and inexpensive pieces of equipment, they are nevertheless indispensable in any chemical plant that handles liquids or gases.

Figure 21.6 illustrates a fluid system containing three components:

1. A flow system, including piping and three heat exchangers
2. A pump
3. A regulating valve

The process shown in Figure 21.6 is described briefly as follows:

A liquid process stream is pumped at a rate of $\dot{v}$ *m³/min from a cooling pond, through a heat exchanger, through a regulating valve and two more heat exchangers connected in series, and returned to the cooling pond. The intake and discharge pipes are at the same elevation. Differential pressure gauges are installed across the pump and the regulating valve. The pump used here has the same characteristics as that shown in Figure 21.4, with a 7-in impeller.*

Conditions shown in Figure 21.6 represent a base case. The mechanical energy balance equation, Equation (21.8), written between points "1" and "2" in Figure 21.6, gives

$$-(\Delta P_{valve} + \Delta P_{pipes} + \Delta P_{exchangers}) = \Delta P_{pump} \qquad (21.14)$$

In Equation (21.14),

ΔP_{valve} represents the pressure drop across the control valve.

ΔP_{pipe} represents the pressure drop due to friction in the piping and pipe fittings between points "1" and "2."

$\Delta P_{exchangers}$ represents the pressure drop due to friction in the three heat exchangers.

ΔP_{pump} represents the pressure increase produced by the pump.

The frictional drop in pressure through the pipe system and the heat exchangers is given by

$$\Delta P_{system} = \Delta P_{pipes} + \Delta P_{exchangers} \qquad (21.15)$$

ΔP_{system} is called the system pressure drop, and Equation (21.14) can be written as

$$-(\Delta P_{valve} + \Delta P_{system}) = \Delta P_{pump} \qquad (21.16)$$

The performance curve for this system, shown in Figure 21.6, consists of a plot of pressure against the liquid flowrate, as was done in Figures 21.4 and 21.5. (In this section pressure units are used instead of units of liquid head.)

The system pressure drop is estimated relative to the base case by Equation (21.13).

$$\Delta P_{2,system} = \Delta P_{1,system}(\dot{v}_2/\dot{v}_1)^2 \qquad (21.17)$$

$$\Delta P_{2,system}(\text{bar}) = 2.15\ (\dot{v}_2/73)^2 \qquad (21.18)$$

This curve has been plotted and labeled as the system curve.

Sufficient information is available to prepare a full performance diagram. In Figure 21.6, for this system, each side of Equation (21.16) is plotted. The pressure produced by the pump, ΔP_{pump}, depends on the volumetric flowrate, $\dot{v}$, as was shown earlier in Figure 21.4. Only the units used to express the pressure and flowrate terms have been changed.

Example 21.6 illustrates how the component terms of the system curve are related to performance parameters.

Example 21.6.

Using the pump curve and the differential pressures provided on Figure 21.6, find for the base case

a. The system pressure drop, ΔP_{system}

b. The volumetric flowrate, $\dot{v}$

Solution

a. For the system pressure drop, ΔP_{system}, from Equation (21.16),

$$-\Delta P_{system} = \Delta P_{pump} + \Delta P_{valve} = 2.9 \text{ bar} - 0.75 \text{ bar} = 2.15 \text{ bar}$$

b. For the volumetric flowrate, $\dot{v}$, see point "b" on Figure 21.6.

$$\text{for } \Delta P_{pump} = 2.9 \text{ bar}, \dot{v} = 73 \text{ m}^3/\text{h}$$

Base-case conditions have been added to Figure 21.6. All pressures for the base case lie along the constant flow line, $\dot{v}$ = 73 m^3/h. Point "c" gives the value for the system pressure drop, ΔP_{system}, and point "b" gives the pressure increase provided by the pump. The difference between these points is the pressure drop over the regulating valve, ΔP_{valve}.

If there were no valve in the line, $\Delta P_{valve} = 0$, the system would operate at point "a," where the pump curve and the system curve cross. It can be seen from Figure 21.6 that as the pressure across the valve increases, the move is toward the left and lower flowrates. The flow becomes zero when the valve is fully closed. Thus, by manipulating the valve setting, the flow of fluid through the system can be altered.

Example 21.7 illustrates obtaining information from a pump and system curve.

Example 21.7.

For the base-case condition shown in Figure 21.6, do the following:

a. Check the pressure drop over the valve against the guideline for control valves (Table 11.8).
b. Determine the percent increase in flow by fully opening the valve.

Solution

a. From guideline, $\Delta P \geq 0.69$ bar (see Table 11.8).
   $\Delta P = 0.75$ bar, and therefore the guideline in Table 11.8 is satisfied.
b. With no valve resistance, $\dot{v}$ = 80 m^3/h.
   The percent increase in flow = [(80 – 73)/73]100 = 9.6%

The increase in flowrate is limited (less than 10% of the base case) by the pump and system, and only modest increases of flowrate are possible for this system.

21.2.3. Reciprocating or Positive Displacement Pumps

Positive displacement pumps perform differently from centrifugal pumps. They are used to achieve higher pressure increases than centrifugal pumps. Figure 21.7 is a drawing of the inner workings of a positive displacement pump. The performance characteristics are represented on Figure 21.8(a). It can be observed that the flowrate through the pump is almost constant over a wide range of pressure increases. One method for regulating the flow through a positive displacement pump is illustrated in Figure 21.8(b). By carefully regulating the flow of the recycle stream, the pressure rise in the pump is controlled. The details were presented in Chapter 18.

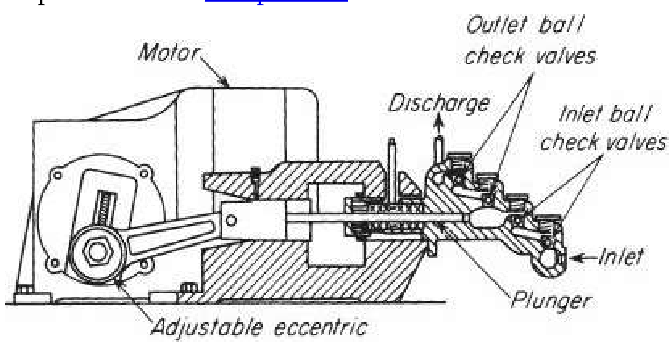

Figure 21.7. Inner Workings of a Positive Displacement Pump (From W. L. McCabe, J. C. Smith, and P. Harriott, *Unit Operations of Chemical Engineering*, 5th ed. New York: McGraw-Hill, 1993. Copyright © 1993 by New York: McGraw-Hill Companies, reproduced with permission of

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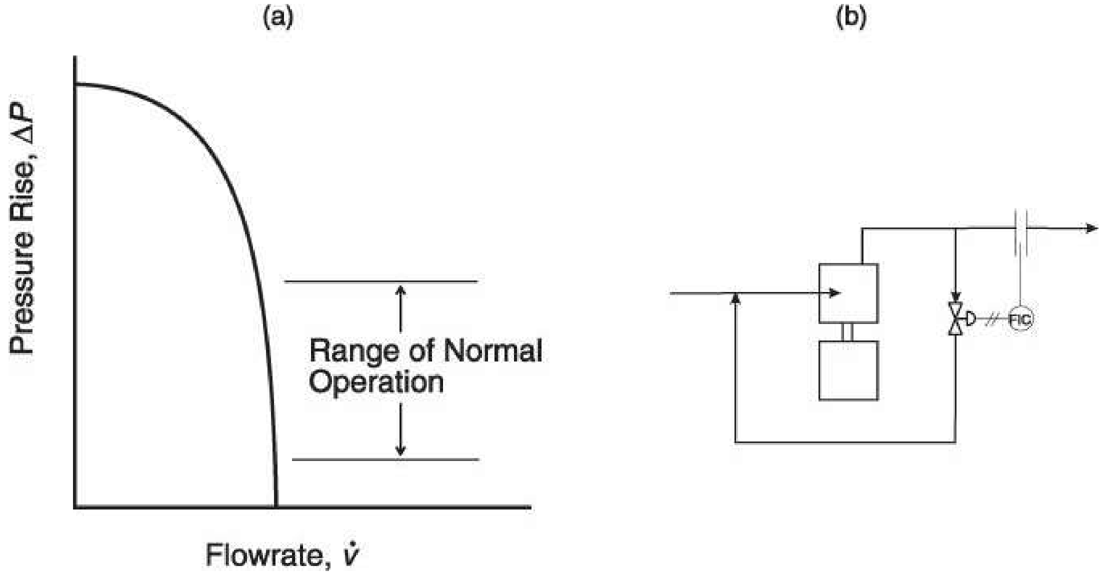

Figure 21.8. Typical Pump Curve for Positive Displacement Pump and Method for Flowrate Regulation

21.2.4. Net Positive Suction Head

There is a significant limitation on pump operation called Net Positive Suction Head (NPSH). Its origin is as follows. Although the effect of a pump is to raise the pressure of a liquid, frictional losses at the entrance to the pump, between the feed (suction) pipe and the internal pump mechanism, cause the liquid pressure to drop upon entering the pump. This means that a minimum pressure exists somewhere within the pump. If the feed liquid is saturated or nearly saturated, the liquid can vaporize upon entering due to the pressure drop. This results in formation of vapor bubbles called **cavitation**. These bubbles rapidly collapse when exposed to the forces created by the pump mechanism. This process usually results in noisy pump operation and, if it occurs for a period of time, will damage the pump. As a consequence, regulating valves are not normally placed in the suction line to a pump.

Pump manufacturers supply NPSH data with a pump. The required NPSH, denoted $NPSH_R$, is a function of the square of velocity, because it is a frictional loss. Figure 21.9 shows an $NPSH_R$ curve, which defines a region of acceptable pump operation. This is specific to a given liquid. Typical $NPSH_R$ values are in the range of 15–30 kPa (2–4 psi) for small pumps and can reach 150 kPa (22 psi) for larger pumps. On Figure 21.9, there are also curves for $NPSH_A$, the available NPSH. The available $NPSH_A$ is defined as

$$NPSH_A = P_{inlet} - P^* \qquad (21.19)$$

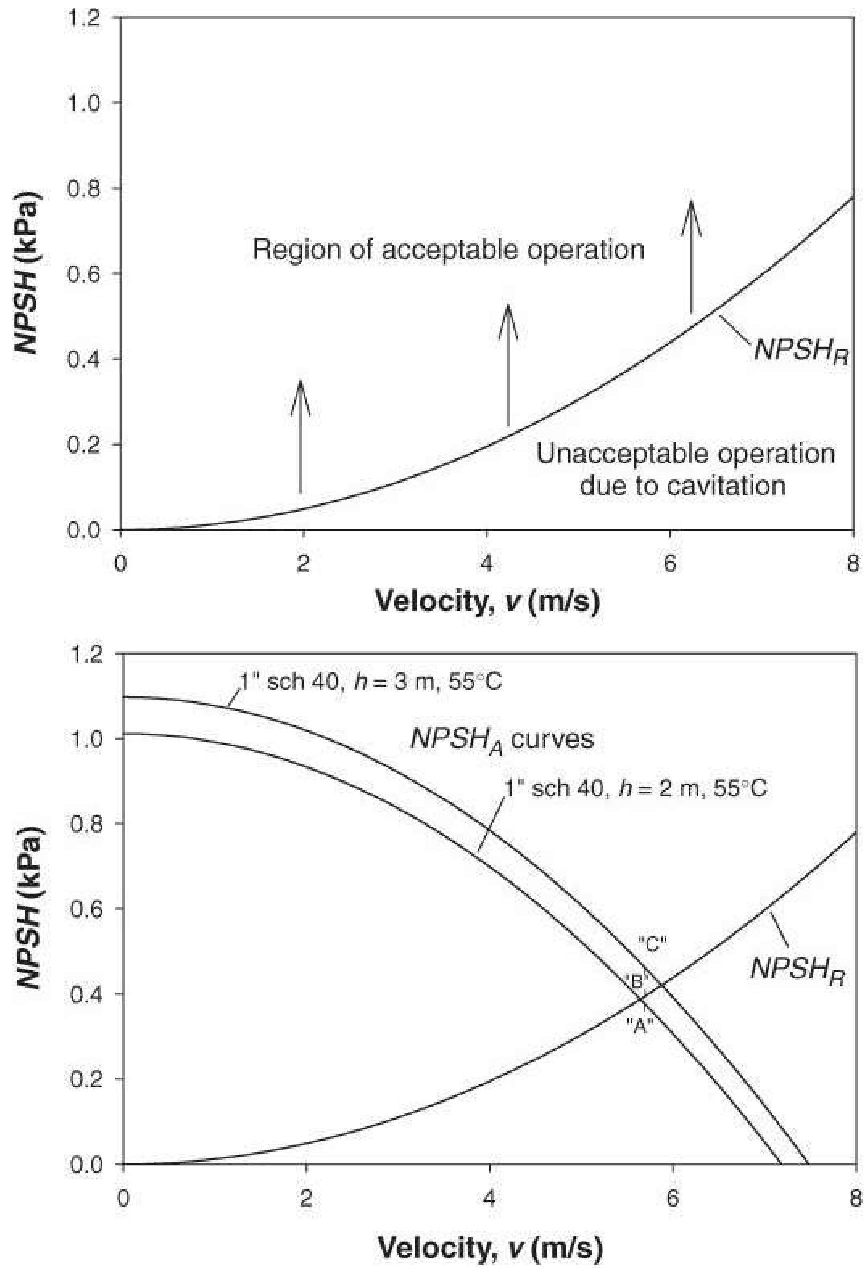

Figure 21.9. $NPSH_A$ and $NPSH_R$ Curves

Equation (21.19) means that the available NPSH ($NPSH_A$) is the difference between the inlet pressure, P_{inlet}, and P^*, which is the vapor pressure (bubble point pressure for a mixture). It is a system curve for the suction side of a pump. It is required that $NPSH_A \geq NPSH_R$ to avoid cavitation. All that remains is to calculate or know the pump inlet conditions in order to determine whether there is enough available NPSH ($NPSH_A$) to equal or exceed the required NPSH ($NPSH_R$).

For example, consider the exit from a distillation column reboiler, which is saturated liquid. If it is necessary to pump this liquid, cavitation could be a problem. A common solution to this problem is to elevate the column above the pump so that the static pressure increase minus any frictional losses between the column and the pump provides the necessary NPSH. This can be done either by elevating the column above ground level using a metal skirt or by placing the pump in a pit below ground level, although pump pits are usually avoided due to safety concerns arising from accumulation of heavy gases in the pit.

In order to quantify NPSH, consider Figure 21.10, in which material in a storage tank is pumped downstream in a chemical process. This scenario is a very common application of the NPSH concept. From the mechanical energy balance, the pressure at the pump inlet can be calculated to be

$$P_{inlet} = P_{tank} + \rho g h - \frac{2\rho f L_{eq} u^2}{D} \qquad (21.20)$$

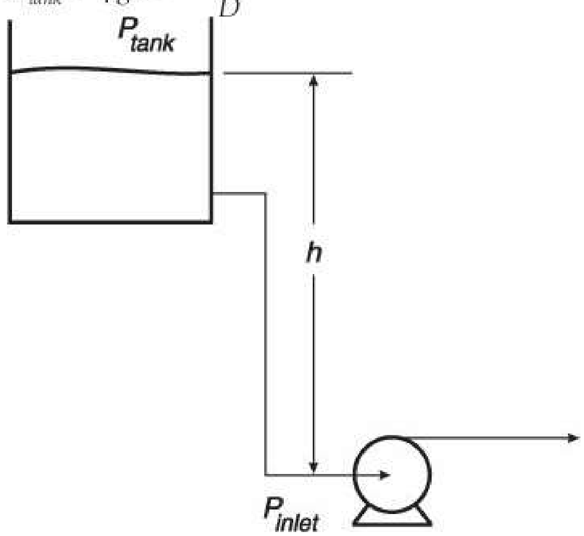

Figure 21.10. Illustration of NPSH for Pumping from Storage Tank

which means that the pump inlet pressure is the tank pressure plus the static pressure minus the frictional losses. Therefore, by substituting Equation (21.20) into Equation (21.19), the resulting expression for $NPSH_A$ is

$$NPSH_A = P_{tank} + \rho g h - \frac{2\rho f L_{eq} u^2}{D} - P^* \qquad (21.21)$$

which is the equation of a concave downward parabola, as illustrated in Figure 21.9.

If there is insufficient $NPSH_A$ for a particular situation, Equation (21.21) suggests methods to increase the $NPSH_A$:

1. Decrease the temperature of the liquid at the pump inlet. This decreases the value of the vapor pressure, P^*, thereby increasing $NPSH_A$.
2. Increase the static head. This is accomplished by increasing the value of h in Equation (21.21), thereby increasing $NPSH_A$. As was said earlier, pumps are most often found at lower elevations than the source of the material they are pumping.
3. Increase the diameter of the suction line (feed pipe to pump). This reduces the velocity and the frictional loss term, thereby increasing $NPSH_A$. It is standard practice to have larger-diameter pipes on the suction side of a pump than on the discharge side.

Example 21.8 illustrates how to do NPSH calculations and one of the methods listed above for increasing $NPSH_A$. The other methods are illustrated in a problem at the end of the chapter.

Example 21.8.

The feed pump (P-101) on Figure 1.5 pumps toluene from a feed tank (V-101) maintained at atmospheric pressure and 55°C. The pump is located 2 m below the liquid level in the tank, and there is 6 m of equivalent pipe length between the tank and the pump. It has been suggested that 1-in schedule-40 commercial steel pipe be used for the suction line. Determine whether this is a suitable choice. If not, suggest methods to avoid pump cavitation.

The following data can be found for toluene: $\ln P^*(\text{bar}) = 10.97 - 4203.06/T(\text{K})$, $\mu = 4.1 \times 10^{-4}$ kg/m s, $\rho = 870$ kg/m^3. For 1-in schedule-40 commercial steel pipe, the roughness factor is about 0.001 and the inside diameter is 0.02664 m. From Table 1.5, the flow of toluene is 10,000 kg/h. Therefore, the velocity of toluene in the pipe can be found to be 5.73 m/s. The Reynolds number is 426,000, and, from a friction factor chart, $f = 0.005$. At 55°C, the vapor pressure is found to be 0.172 bar.

From Equation 21.21,

$$NPSH_A = 1.01325 \text{ bar} + 870(9.81)(2)(10^{-5}) \text{ bar}$$

$$- 2(870)(0.005)(6)(5.73)^2(10^{-5})/(0.02664) \text{ bar} - 0.172 \text{ bar}$$

$$NPSH_A = 0.37 \text{ bar}$$

This is shown as point "A" on Figure 21.9. At the calculated velocity, Figure 21.9 shows that $NPSH_R$ is 0.40 bar, point "B." Therefore, there is insufficient $NPSH_A$.

One method for increasing $NPSH_A$ is to increase the height of liquid in the tank. If the height of liquid in the tank is 3 m, with the original 1-in schedule-40 pipe at the original temperature, $NPSH_A = 0.445$ bar. This is shown as point "C" on Figure 21.9.

21.2.5. Compressors

The performance of centrifugal compressors is somewhat analogous to that of centrifugal pumps. There is a characteristic performance curve, supplied by the manufacturer, that defines how the outlet pressure varies with flowrate. However, compressor behavior is far more complex than that for pumps because the fluid is compressible.

Figure 21.11 shows the performance curve for a centrifugal compressor. It is immediately observed that the *y*-axis is the ratio of the outlet pressure to inlet pressure. This is in contrast to pump curves, which have the difference between these two values. Curves for two different rotation speeds are shown. As with pump curves, curves for power and efficiency are often included but are not shown here. Unlike most pumps, the speed is often varied continuously to control the flowrate. This is because the higher power required in a compressor makes it economical to avoid throttling the outlet as in a centrifugal pump.

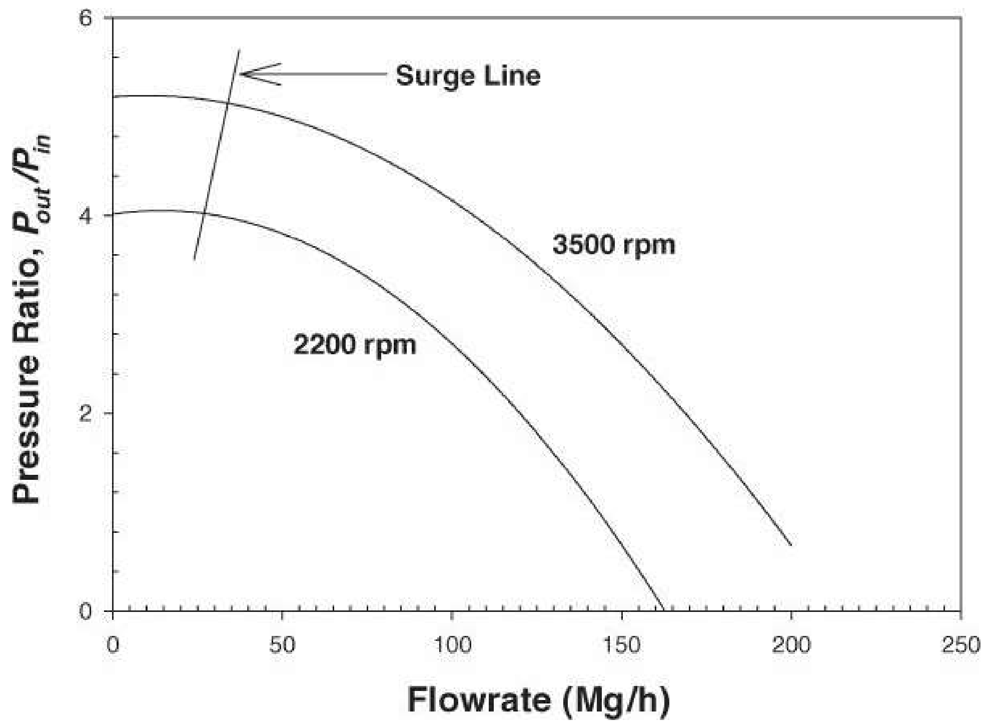

Figure 21.11. Centrifugal Compressor Curves

Centrifugal compressor curves are read just like pump curves. At a given flowrate and rpm there is one pressure ratio. The pressure ratio decreases as flowrate increases. A unique feature of compressor behavior occurs at low flowrates. It is observed that the pressure ratio increases with decreasing flowrate, reaches a maximum, and then decreases with decreasing flowrate. The locus of maxima is called the surge line. For safety reasons, compressors are operated to the right of the **surge line**. The surge line is significant for the following reason. Imagine that starting at a high flowrate and the flowrate is lowered continuously, causing a higher outlet pressure. At some point, the surge line is crossed, lowering the pressure ratio. This means that downstream fluid is at a higher pressure than upstream fluid, causing a backflow. These flow irregularities can severely damage the compressor mechanism, even causing the compressor to vibrate or surge (hence the origin of the term). Severe surging has been known to cause compressors to become detached from the supports keeping them stationary and literally to fly apart, causing great damage. Therefore, the surge line is considered a limiting operating condition, below which operation is prohibited.

Positive displacement compressors also exist and are used to compress low volumes to high pressures. **Centrifugal compressors** are used to compress higher volumes to moderate pressures and are often staged in order to obtain higher pressures. Figure 21.12 illustrates the inner workings of a compressor.

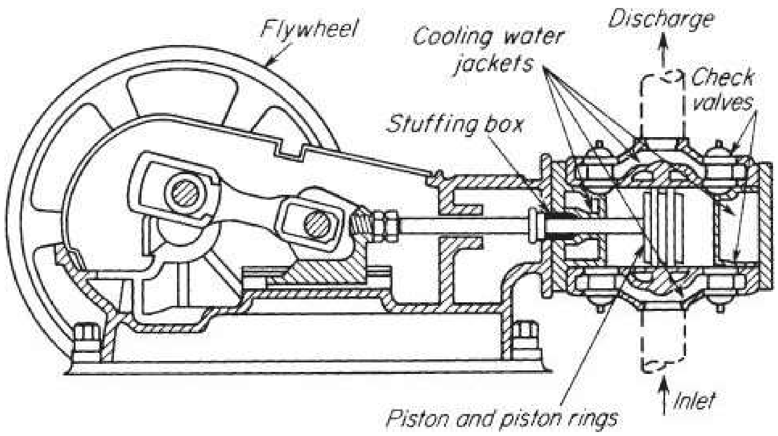

Figure 21.12. Inner Workings of a Positive Displacement Compressor (From W. L. McCabe, J. C. Smith, and P. Harriott, *Unit Operations of Chemical Engineering*, 5th ed. New York: McGraw-Hill, 1993. Copyright © 1993 by the McGraw-Hill Companies, reproduced with permission.)

Problems requiring reading compressor curves are given at the end of the chapter.

21.3. Application to Separation Problems

21.3.1. Separations with Mass Separating Agents

Multistage equilibrium separations involve simultaneous solution of material balances and equilibrium relationships for each equilibrium stage. Therefore, there is no simple, closed-form relationship that describes the behavior of these systems for all situations. There are qualitative relationships that are applicable to almost all situations, and these were included in Table 20.1. The key relationships are that a better separation is usually achieved by increasing the number of stages or by increasing solvent flowrate.

Similarly, continuous differential equilibrium separations, which involve simultaneous solution of material balances and mass transfer relationships, do not yield a closed-form solution either. The key qualitative relationships are that better separation is usually achieved by increasing column height or by increasing flowrate.

For the specific assumptions of dilute solutions and a linear equilibrium relationship that results in approximately constant stream flows, an analytical solution is possible for the above situations. For staged separations, the situation is shown in Figure 21.13(a). The result is

$$\frac{y_{A,out} - y^*_{A,out}}{y_{A,in} - y^*_{A,out}} = \frac{1 - A}{1 - A^{N+1}} \qquad (21.22)$$

where $A = (L/mG)$ is called the absorption factor, N is the number of equilibrium stages, y_A is the mole fraction of the solute in the gas phase, L and G are the molar flowrates of each stream, m is the equilibrium relationship ($y = mx$), and $y^*_{A,out} = mx_{A,in}$. Equation (21.22) is known as the Kremser equation and is a key relationship for multistage equilibrium separations obeying the assumptions listed above. It describes transport of solute, A, from the G phase to the L phase. For the reverse situation, a similar equation is derived in which the left side of Equation (21.22) involves x_A, the mole fraction in the liquid phase, and the right side involves $S = (mG/L)$. The term $y^*_{A,out} = mx_{A,in}$ is replaced by $x^*_{A,out} = {}_{A,in}/m$. If the separation involves phases that are not gas or liquid, Equation (21.22) may still be used by defining L and G appropriately. Figure 21.14 is a plot of Equation (21.22) and contains the performance relationship between key variables for multistage equilibrium separations following the assumptions listed above. Performance curves for a specific staged separation can be generated from the information in Figure 21.14. This is illustrated in Example 21.10. It should be noted that tray performance issues such as flooding, which are specific to a particular column design, are not predicted by the Kremser relationship.

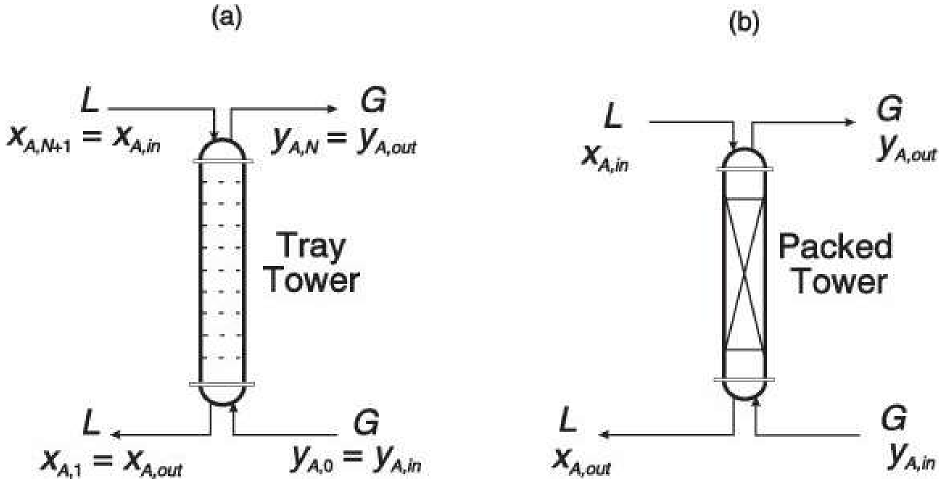

Figure 21.13. Tray and Packed Absorbers

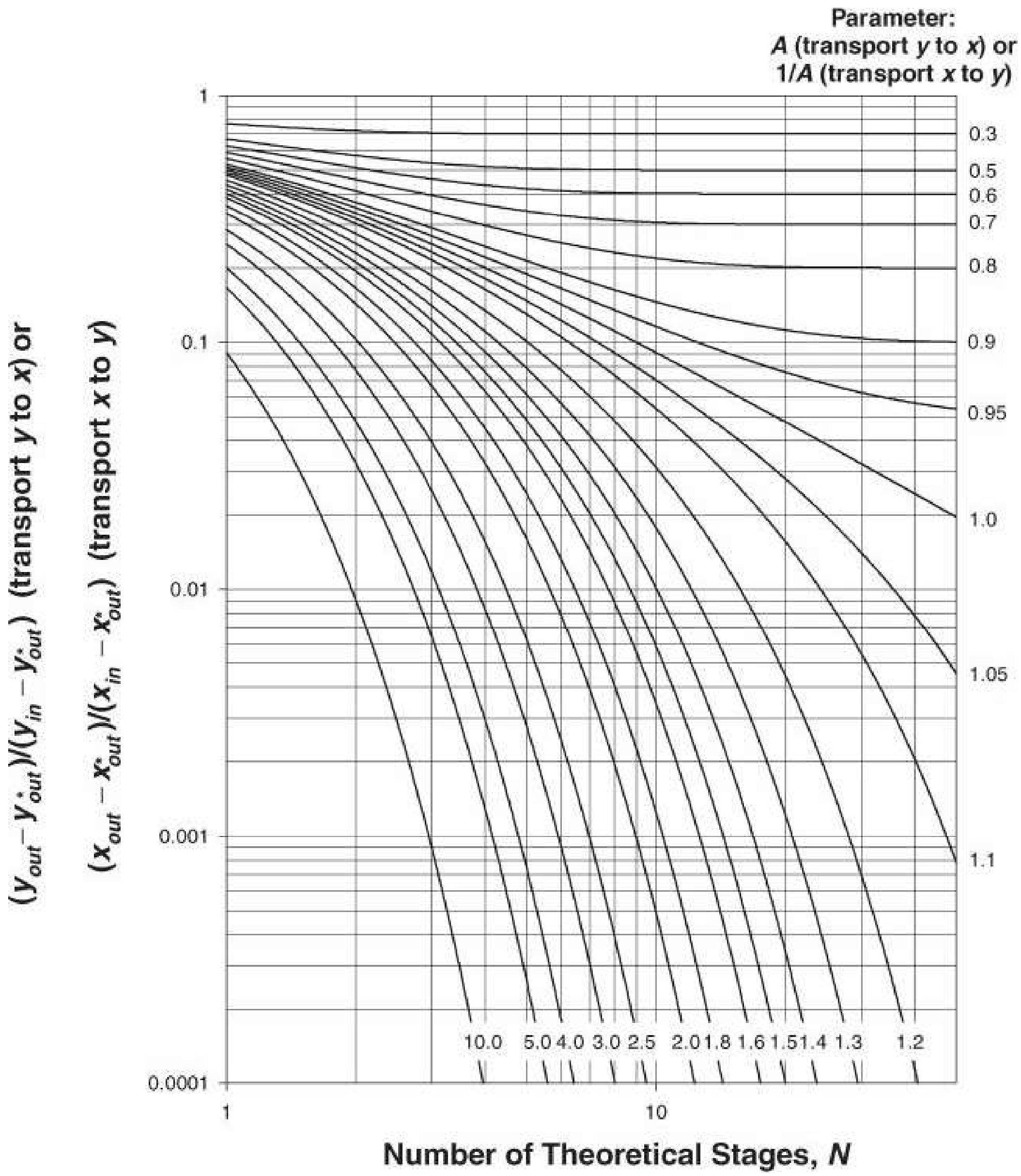

Figure 21.14. Plot of Kremser Equation, Number of Theoretical Stages for Countercurrent Operation, Henry's Law Equilibrium, and Constant *A* or 1/*A*

Even though Equation (21.22) and Figure 21.14 are only valid subject to the assumptions listed above, the qualitative performance of all staged separations can be understood from these relationships. For example, as illustrated in Figure 21.14, if a line of constant A (if $A > 1$) is followed for an increasing number of stages, N, it is seen that the separation continues to improve (lower value of y-axis). For $A < 1$, a best separation is approached asymptotically. If a vertical line of constant N is followed, it is seen that as the separation improves, the value of A increases. A increases if L increases or if G decreases, meaning a larger solvent to feed ratio, or if m decreases, meaning equilibrium more in favor of the L phase.

A similar relationship to the Kremser equation exists for continuous differential separations (packed beds), subject to the same assumptions. The physical situation is illustrated in Figure 21.13(b). This relationship, known as the Colburn equation, is

$$\frac{y_{A,in} - y^*_{A,out}}{y_{A,out} - y^*_{A,out}} = \frac{e^{N_{tOG}[1-(1/A)]} - (1/A)}{1 - (1/A)} \tag{21.23}$$

where N_{tOG} is the number of overall gas-phase transfer units. Equation (21.23) is for absorption, that is, transfer into the L phase. For the reverse direction of transport, the same changes are made as in the Kremser equation, with N_{tOL}, the number of overall liquid-phase transfer units, replacing N_{tOG}. Equation (21.23) is plotted in Figure 21.15. Performance curves for a specific staged separation can be generated from the information in Figure 21.15. This is the subject of a problem at the end of the chapter. Because increasing the number of transfer units increases column height, the same qualitative understanding applicable to all systems is gleaned from Figure 21.15 as was described above. For increasing N_{tOG} (which means increased column height) at constant A, a better separation is observed (for $A > 1$). For $A < 1$, a best separation is approached asymptotically. Similarly, at constant column height (constant N_{tOG}), increasing A results in better separation.

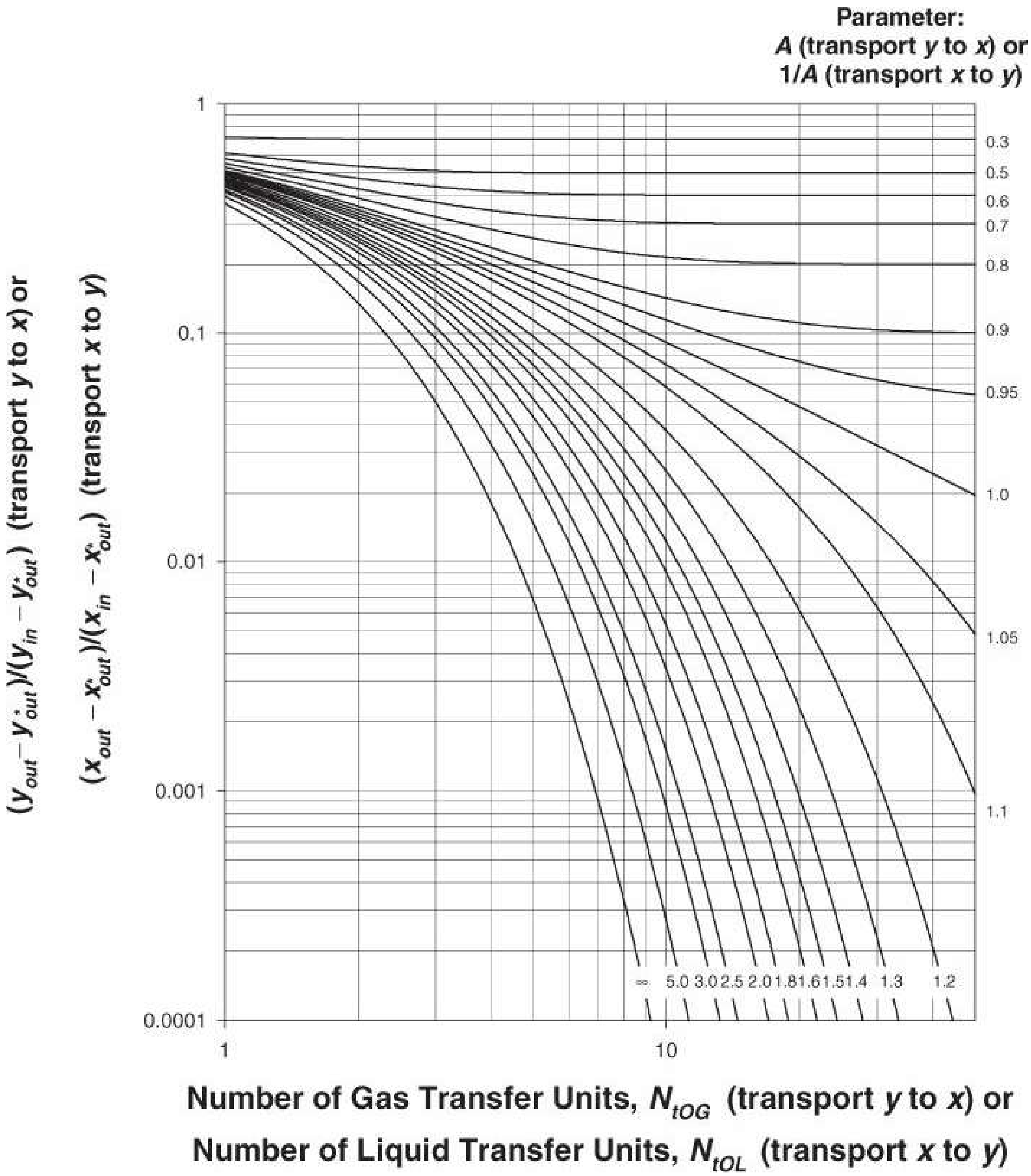

Figure 21.15. Plot of Colburn Equation, Number of Transfer Units for Countercurrent Operation, Henry's Law Equilibrium, and Constant *A* or 1/*A*

As with the Kremser equation, the Colburn equation does not describe flooding behavior of a packed bed, which is specific to a particular column design. Flooding occurs when the vapor velocity upward through the column is so great that liquid is prevented from flowing downward. When a column is designed, the diameter is chosen so that the vapor velocity is below the flooding limit (typically 75%–80% of the limit). However, if the vapor velocity is increased, flooding can occur. Because flooding is specific to a particular column, it cannot be illustrated on the general performance curve. Care must be taken not to use the Kremser or Colburn graphs to obtain a result that will cause flooding in a particular column or to recommend operation in the flooding zone.

Examples 21.9 and 21.10 illustrate the use of the Kremser and Colburn equations.

Example 21.9.

A tray scrubber with eight equilibrium stages is currently operating to reduce the acetone concentration in 40 kmol/h of air from a mole fraction of 0.02 to 0.001. The acetone is absorbed into a 20 kmol/h water stream, and it can be assumed that the water stream enters acetone free. Due to a process upset, it is necessary to increase the flow of air by 10%. Under the new operating conditions, what is the new outlet mole fraction of acetone in air?

The use of Figure 21.14 for this problem is illustrated in Figure E21.9. For the design case, $y_{A,in}$ = 0.02, $y_{A,out}$ = 0.001, and $x_{A,in}$ = 0, which means that $y^*_{A,out} = 0$. Therefore, the *y*-axis is 0.05. Because *N* = 8, the design case has a value of *A* = 1.2. This is shown as point "a" on Figure E21.9. If the flow of air (*G*) is increased by 10%, then the new value of *A* = 1.08. Following a vertical line (constant number of stages) from the original point to a value of *A* = 1.08 (point "b") yields a *y*-axis value of 0.08. Hence, the new value of $y_{A,out}$ = 0.0018.

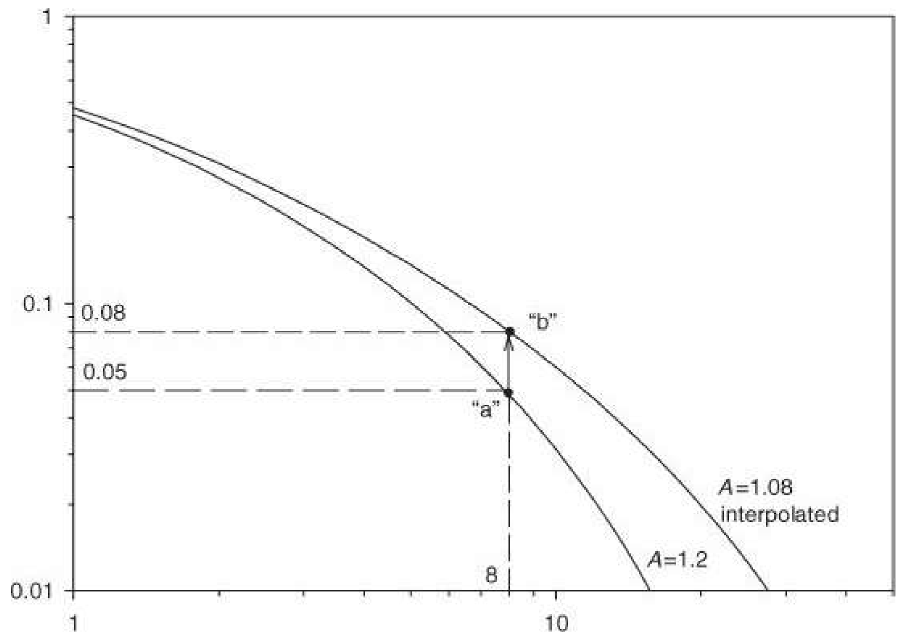

Figure E21.9. Use of Kremser Graph to Solve Example 21.9

Example 21.10.

For the situation described in Example 21.9, the inlet air flowrate and the inlet mole fraction of acetone in air may vary during process operation. Prepare a set of performance curves for the liquid rate necessary to maintain the outlet acetone mole fraction for a range of inlet acetone mole fractions. There should be curves for five different values of the inlet gas rate, including the original case and the following percentages of the original case: 80, 90, 110, and 120. The temperature and pressure in the scrubber are assumed to remain constant.

The result is shown on Figure E21.10. The curves are generated as follows. From the original operating point, the value of *m* can be determined to be 0.417. By moving vertically on a line of constant number of equilibrium stages (eight in this case), values for the *y*-axis and *A* can be tabulated. The *y*-axis value is easily converted into $y_{A,in}$ because $y_{A,out}$ is known and constant. Then, for different values of *G*, values of *L* are obtained because *m* is known and constant.

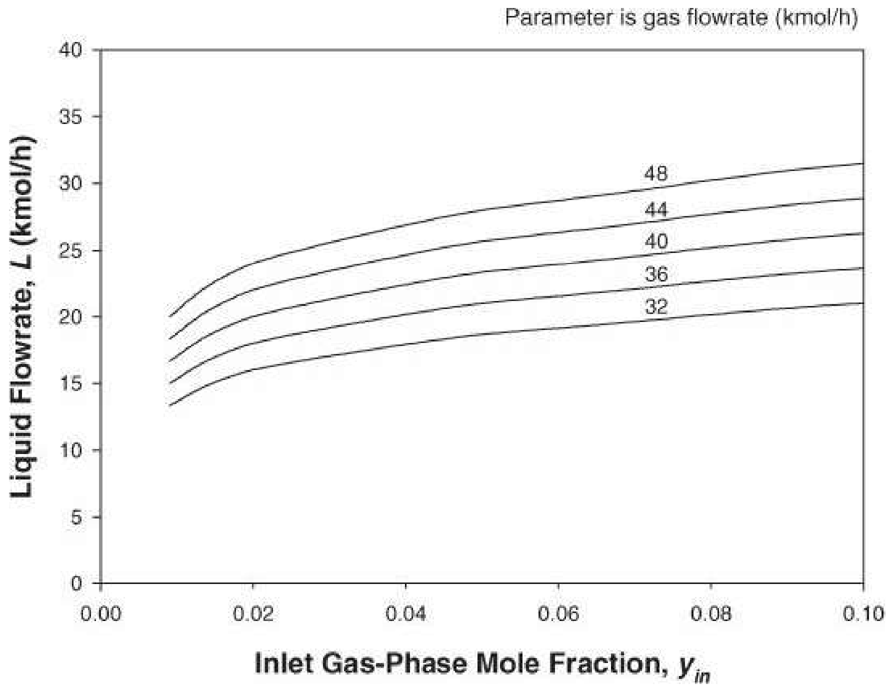

Figure E21.10. Performance Curves for Example 21.10

21.3.2. Distillation

For distillation, there is no universal set of performance curves as for multistage equilibrium separations involving mass separating agents. However, for a given situation, performance curves can be generated. This section demonstrates how this is done.

Figure 21.16 illustrates a multistage distillation system for separating benzene from a mixture of benzene and toluene. Figure 21.17 is a sketch of a typical distillation column. This problem is modified from Bailie and Shaeiwitz [1].

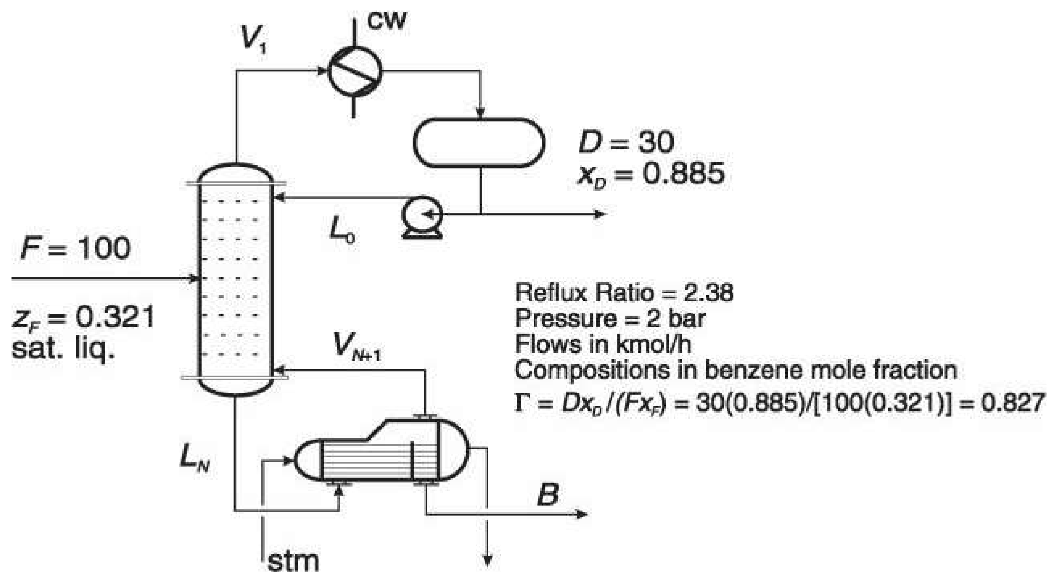

Figure 21.16. Plant Section for Distillation of Benzene from Toluene

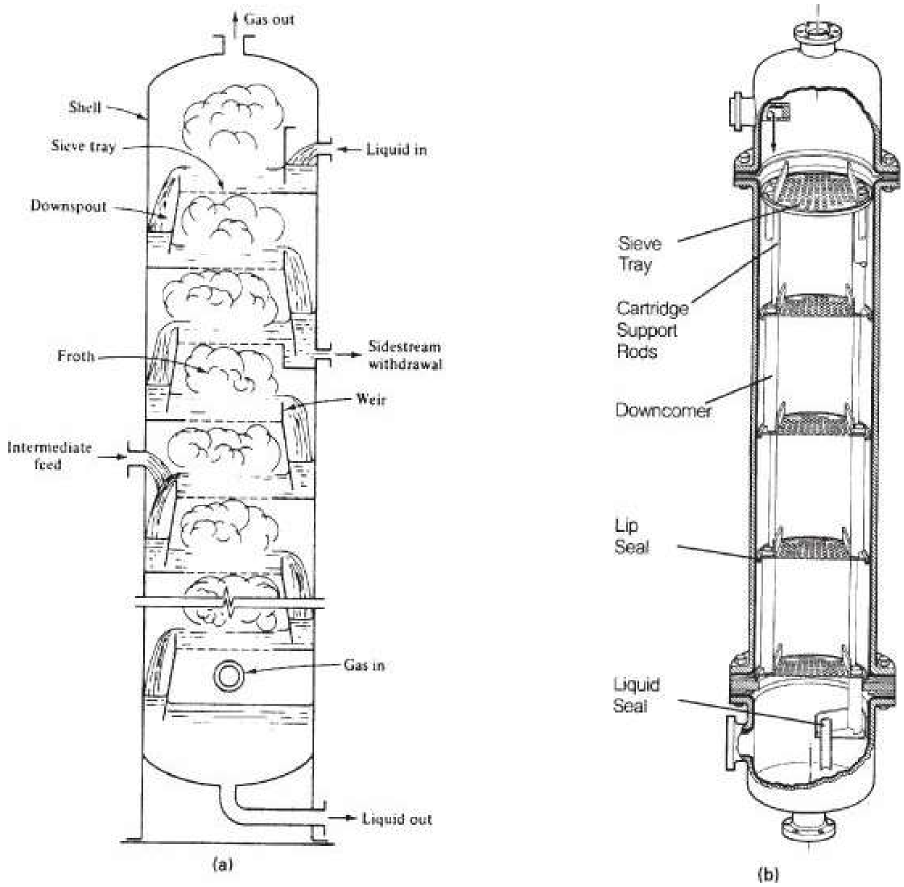

Figure 21.17. Details of Internal Construction of a Distillation Column (From S. Walas, *Chemical Process Equipment: Selection and Design.* Stoneham, MA: Butterworth, 1988. Reprinted with permission.)

Component separation takes place within the tower. Considering the tower alone, there are three input and two output streams. Separation is achieved by the transfer of components between the two phases. For an operating plant, the size of the tower, the number and type of trays (or height of packing) in the tower, and other tower attributes are fixed. The output streams are established once the input flow streams to the tower are set.

In this section several performance diagrams will be developed for the distillation process shown in Figure 21.16. The tower is the separation unit, and the two phases needed for the separation are provided by the condenser and reboiler.

In a typical distillation system, two recycle streams are returned to the tower. A condenser is added at the top of the column, and a fraction of the overhead vapor, V_1, is condensed to form a liquid recycle, L_0. This provides the liquid phase needed in the tower. The remaining fraction is the overhead product, *D*. A vaporizer (reboiler) is added to the bottom of the column, and a portion of the bottom liquid, L_N, is vaporized and recycled to the tower as stream V_{N+1}. This provides the vapor phase needed in the tower.

Figure 21.16 provides some information on the distillation process taken from an operating plant.

> *Product benzene is separated from a benzene-toluene mixture. The feed stream,* F = *100 kmol/h, is a saturated liquid with composition* z_F = *0.321 (mole fraction benzene) at a pressure*

P = 2 bar. The top product consists of a distillate stream, D = 30 kmol/h, with composition x_D *= 0.885 and a corresponding benzene recovery,* Γ *= 0.827. The tower has a diameter of 0.83 m and contains 20 sieve plates with a 0.61 m plate spacing. Feed is introduced on Plate 13, and the tower operates at a reflux ratio of 2.38. No information is provided on the partial reboiler and total condenser. From a benzene balance,* B *= 70 kmol/h and* x_B *= 0.079.*

Figure 21.18 provides a McCabe-Thiele diagram for the separation given in Figure 21.16. This was constructed from the equilibrium curve and by matching the concentrations, $x_D = 0.885$, $z_F = 0.321$, and $x_B = 0.079$, the reflux ratio, $R = 2.38$, and the condition of saturated liquid feed with plant operating conditions. This McCabe-Thiele construction yielded 13 theoretical stages (12 trays plus a reboiler). This means that the average plate efficiency was 60% (100[No. of theoretical plates/No. of actual plates] = 100(12/20). The 60% plate efficiency is within the range found for columns separating benzene from toluene. The feed is added on theoretical plate number 8 (13[0.60] ≈ 8). (*Note:* The number 13 appearing in the previous calculation refers to the actual feed location and not to the number of stages.)

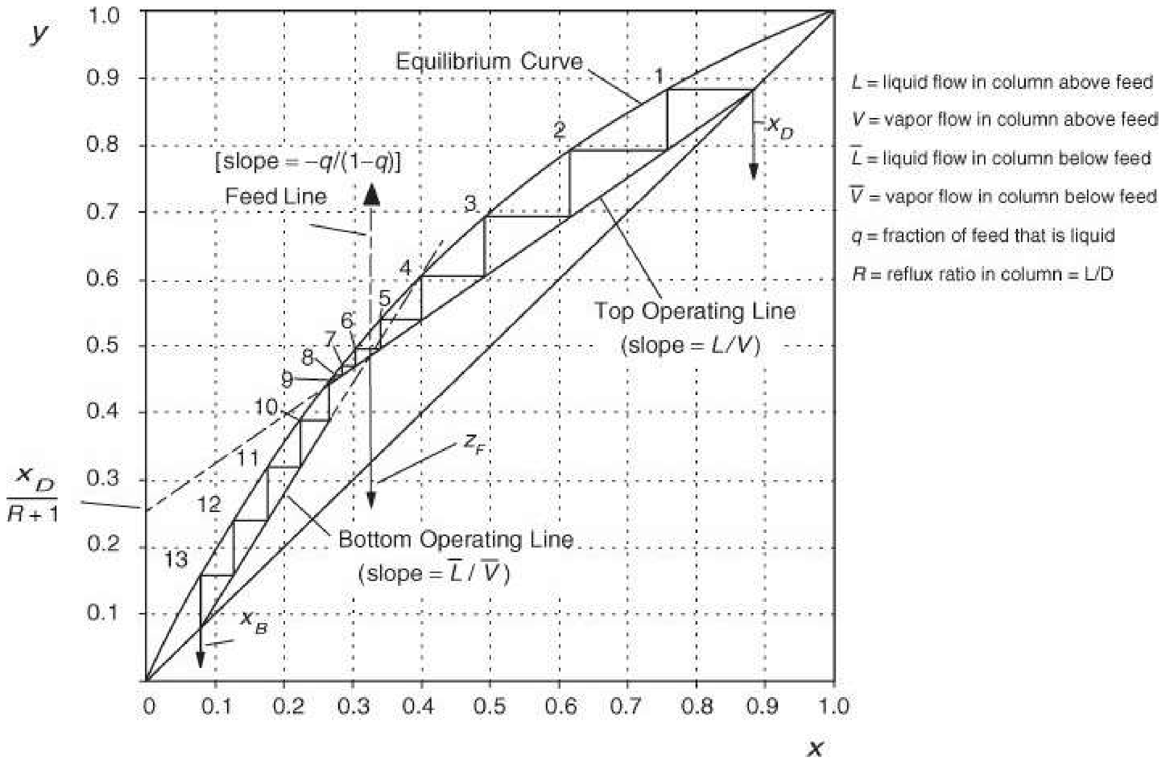

Figure 21.18. McCabe-Thiele Diagram for Benzene Separation: Base Case

The McCabe-Thiele method for evaluating theoretical stages is limited by the following assumptions and constraints:

1. It applies only to binary systems.
2. It assumes “constant molal overflow” and in most situations does not satisfy an overall energy balance.
3. It requires graphical trial-and-error solutions to solve performance problems.

Even though the speed of computation using the McCabe-Thiele method is far slower than for process simulators, it nevertheless remains an important analytical tool. Its graphical representation promotes a clarity of understanding that leads to valuable insights. This is illustrated by using the McCabe-

Thiele diagram to glean some critical information required to construct a performance curve for the distillation process in Figure 21.18.

Upon studying Figure 21.18 the following factors are revealed:

1. Feed is not being introduced at the optimum location in the column. The optimal location is on Plate 10 (6/0.60 ≈ 10) and not Plate 13 as currently used.
2. Separation steps near the feed plate are small.

For the current conditions and feed plate location, there are more than the optimum number of stages in the rectifying (top) section. This suggests that the unit may have been designed to process a lower concentration of feed material or to produce a higher-concentration distillate stream than is required currently.

Increasing the slope of the top operating line, L/V, moves the operating lines away from the equilibrium line. This increases the separation of each stage (increases the **step size**). In the special case when the distillate concentration remains the same, the concentration of the more volatile material (in this case, benzene) in the bottom stream will be lowered.

From an inspection of the McCabe-Thiele diagram, the tower appears to be oversized for the present separation. It could process a lower quality of feed (lower benzene concentration), produce a higher-quality product, and improve benzene recovery. These are points that may deserve further investigation and could be overlooked easily without the graphical representation provided by the McCabe-Thiele diagram.

The McCabe-Thiele analysis is difficult to utilize for developing all the information needed in a performance problem. It requires a graphical trial-and-error solution to match the number of stages to a set of values for z_F, x_D, and x_B. Fortunately, modern computer simulators can provide rigorous solutions for a large variety of separators, including distillation towers. The information needed to construct the performance diagrams for the distillation column in this chapter was obtained from a CHEMCAD simulation. As a base case, a column with 13 theoretical stages (fed on Tray 8) was selected. In the preliminary evaluation, it was assumed that the tray efficiency remained constant. This assumption will be revisited in the discussion of the limitations of these performance curves.

There are three input streams entering the tower shown in Figure 21.16. The flowrate and concentration of these streams, coupled with the performance of the equipment, establish the tower output. The remaining streams in the distillation system can be calculated once these streams are known.

The results of a computer simulation for the base-case conditions are shown in Table 20.1.

Table 20.1. Results of Computer Simulation for Base-Case Conditions

	Flowrate (kmol/h)	Mole Fractions	Concentration
Inputs			
F	100.0	z_F	0.321
Outputs from Simulation			
L_0	71.5	x_0	0.885
V_{N+1}	96.4	y_{N+1}	0.079
V_1	101.5	y_1	0.885
L_N	166.4	x_N	0.079

The value for the vapor flowrate changed from 101.5 kmol/h to 96.4 kmol/h moving from the top to the bottom of the column. Even for this “ideal” separation, the assumption of constant molal overflow is not satisfied.

The process simulator provides a rigorous solution by carrying out material balances, energy balances, and equilibrium calculations over each stage. The simulator provides tray-by-tray results that include the composition of liquid and vapor on each tray, temperature, pressure, and K-values for each component on each tray, along with transport properties of both phases. Using data from this simulation, performance diagrams may be obtained. The most important variables are the tower inputs. They cause the outputs to change. Selection of which performance diagrams to prepare depends on the problem being considered. This is illustrated in the following case study.

> You are the engineer in charge of the toluene-benzene distillation section in a chemical plant. You have just met with your supervisor and have been informed that, in the future, the feed concentration to this unit will no longer be constant but will vary between 25% and 40% benzene. You have also been told that it is important to maintain a constant distillate flow, D, and concentration, x_D, from this distillation unit to the downstream process.
>
> At this meeting, possibilities for replacing parts of this process, replacing the whole system, and installing a storage system to blend feed and/or product distillate, or a combination of these alternatives, were discussed. The object is to maintain the distillate flow and concentration.
>
> To make a responsible decision on this matter, it will be necessary to assess the effect that changes in feed concentration will have on the operation of the distillation process. Before making a decision, background material will be needed. You have been told to provide a performance diagram that shows the effects of changes in feed concentration on the reflux flowrate. For a preliminary assessment, you are to assume that the reboiler and condenser can meet any new demands. If they are found inadequate, they will be replaced or modified.
>
> In addition, questions have been raised regarding the low benzene recovery in the current operation, and you have been told to consider the impact of using higher benzene recoveries.

For the situation described above, the important variables are the feed concentration, x_F, and the benzene recovery, Γ. The information needed to respond to the above request can be obtained by running simulations at different feed concentrations and reflux flowrates to determine the benzene recovery for each pair of conditions. Reflux is chosen as the dependent variable here because the flow of the reflux stream is the most common way to regulate the performance of a distillation column.

Figure 21.19 gives the performance curves for the reflux, L_0. The performance diagram shows the feed concentration, z_F, on the x-axis with the benzene recovery, Γ, as a parameter. The base case is identified as point “a” on the diagram.

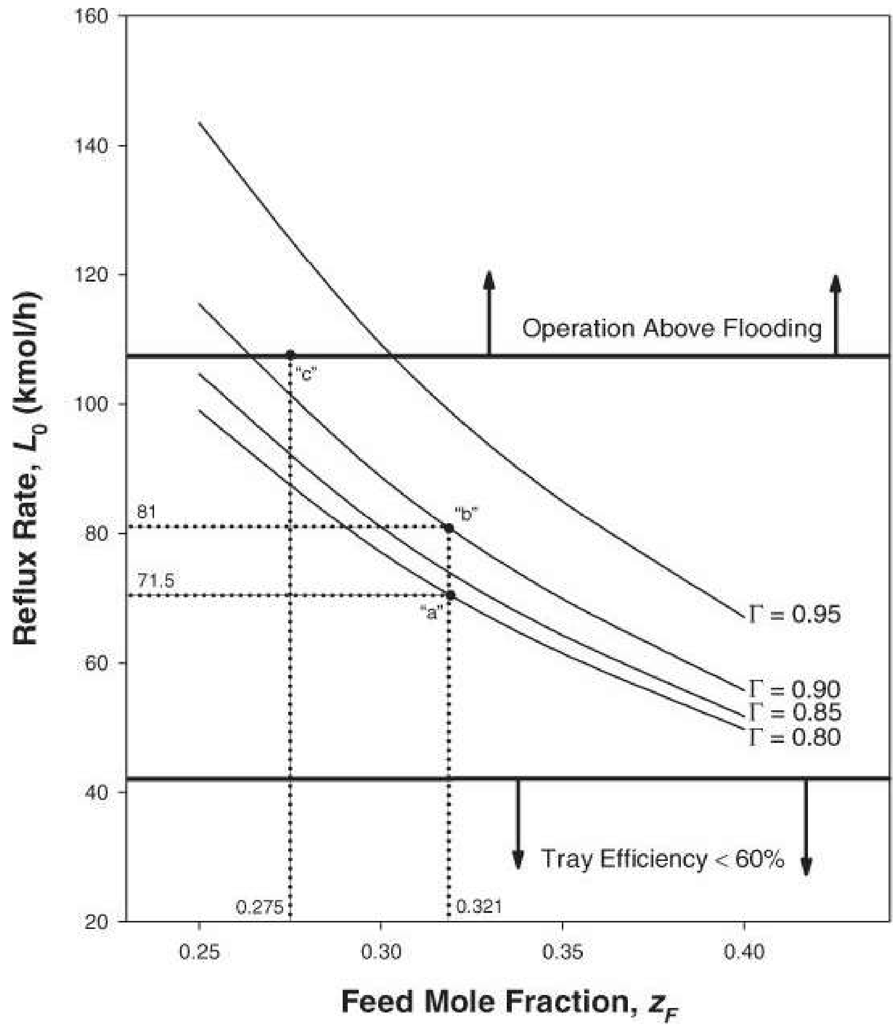

Figure 21.19. Performance Curve for Distillation Tower

The performance curves presented show the following trends:

1. **For constant recovery, Γ:** As z_F increases, the reflux decreases. The rate of decrease becomes less as z_F increases.
2. **For a constant feed concentration, z_F:** As Γ increases, the reflux increases. The rate of increase becomes greater as Γ increases.

The performance diagram given in Figure 21.19 is specific to the existing tower with fixed feed location, saturated liquid feed, and desired distillate conditions. The conditions or variables that are fixed in this problem are as follows:

D = 30 kmol/h

x_D = 0.885 mole fraction benzene

20 trays (12 theoretical equilibrium stages plus a reboiler)

Feed plate = 13 (theoretical equilibrium Stage 8)

Diameter of tower = 0.83 m

Tray spacing = 0.61 m

Overall tray efficiency = 60%

The independent input variables are the feed concentration, z_F, and the benzene recovery, Γ. One important point to note is that, as in all other performance problems, the parameters associated with the equipment are fixed. A second point to note is that with the information in Figure 21.19, it is possible to work backward and estimate the desired reflux flowrate, for example, that would be required to obtain the desired distillate conditions if the feed concentration changed from 0.321 to 0.295. The utility of the performance curves is best illustrated in Example 21.11. Additional problems involving construction and use of performance curves for distillation are given at the end of the chapter.

Example 21.11.

Using the performance diagram, Figure 21.19, find the changes to the tower input needed to achieve a recovery of 90% for the current feed concentration, $x_F = 0.321$.

New Feed Rate, F: $F = x_D D/(z_F \Gamma) = 0.885(30)/[(0.321)(0.9)] = 91.9$ kmol/h

New Reflux, L_0: From Figure 21.19 new feed rate (see point "b") $L_0 = 81$ kmol/h

---

Example 21.11 showed that the distillation system could be operated at a recovery, Γ = 0.9. This reduces the feed that can be processed by 8.1%.

Installed equipment imposes operating limitations that restrict the range over which the performance diagrams can be used. Equipment manufacturers often provide technical information that gives the limitations of the equipment. This information may include, but is not limited to, operating temperature, pressure, installation instructions, and operating parameters such as plate efficiencies and flooding and weeping velocities. Thus this specific information adds additional constraints on the range of operations that may be considered.

From the technical information provided with the sieve trays used in the distillation tower in Figure 21.19, the following information was obtained:

1. Flooding gas velocity, $u_f = 1.07$ m/s
2. Weeping gas velocity, $u_w = 0.35$ m/s
3. Tray efficiency, $\epsilon = 0.60$ for $2.6 > u[\text{m/s}](\rho_v[\text{kg/m}^3])^{0.5} > 1.2$

Based on this information, two regions on Figure 21.19 are shown. These are "forbidden" or infeasible regions, in which the performance curves are not valid because of the limitations of the specific equipment or the assumptions made (e.g., $\epsilon = 0.60$). For any change to be made, the new condition must be checked to ensure that the new operating point lies within the feasible region for all the performance curves.

The flooding velocity limit is also shown on Figure 21.19. The limit related to weeping lies below that for tray efficiency; that is, the tray efficiency drops below 60% well before the weeping condition is reached. The constraint of operating below a tray efficiency of 60% is shown as the lower infeasible region in Figure 21.19.

Example 21.12 shows how the flooding limit of the equipment, illustrated on the performance graph, identifies another equipment limitation.

Example 21.12.

Find the maximum recovery possible from the distillation equipment for a feed concentration, z_F =

0.275.

Point “c” on Figure 21.19 provides an estimate of the recovery, $\Gamma = 0.91$.

The constraints associated with the equipment must be included in any analysis of performance.

In the preliminary analysis, it was assumed that the tray efficiency was 60% and the reboiler and condenser were adequate. From the manufacturer’s data, it was confirmed that the efficiencies were constant over the operating range given above. The limits on the reboiler and the condenser have not been considered in the above analysis and must be checked against the performance limits for these pieces of equipment. In addition, the curves obtained assume that the feed location would not change. However, by using the optimum location, as the feed concentration changes, the performance of the tower could be improved. The use of alternative feed locations requires that feed nozzles be present at different trays. Thus, once again, the constraints of the existing equipment dictate whether this option should be considered further.

21.4. Summary

In this chapter how to construct simple performance diagrams for some common individual pieces of equipment was presented. These performance diagrams show how a given piece of equipment responds to changes in input flows. They also can be used to predict what changes in input variables would be required in order to obtain a desired output condition. In some instances, as with valves, pumps, and compressors, performance curves are provided by equipment manufacturers. In other instances, as with the Kremser or Colburn equation, universal performance curves and equations exist subject to specific assumptions. In still other instances, as with heat exchangers or distillation columns, performance curves must be constructed by modeling equipment behavior.

In all real chemical processes, the final element that makes changes in flowrates possible is the regulating valve. Without such valves, it would be impossible for a process to adjust for unforeseen changes in operating conditions. In addition, it would be equally impossible to manipulate a process to give new desired outputs. Regulating valves are relatively simple and inexpensive pieces of equipment, and yet they are absolutely essential in the day-to-day operation of a chemical process. The use of regulating valves was illustrated for a simple flow system including a pump, piping system, process equipment, and a valve.

What You Should Have Learned

- How performance relates the output to the input when the equipment is fixed (already exists)
- The basic analytical method for doing performance calculations involving ratios
- That the constraints of chemical process equipment must be included in this analysis
- How to represent the performance of typical chemical process equipment graphically

Reference

1. Bailie, R. C., and J. A. Shaeiwitz, “Performance Problems,” *Chem. Eng. Edu.* 28 (1994): 198–203.

Short Answer Questions

1. What is a pump curve? Sketch a typical pump curve, making sure that both axes are clearly labeled. On the same sketch, show a typical system curve. Indicate the typical region of good operating practice. Explain.
2. Comment on the following statement: *For a two-pump system, it is always best to run the two pumps in series rather than in parallel because greater scale-up will be possible.*
3. Explain the meaning of the intersection of the two curves on the diagrams in Figure P21.3.

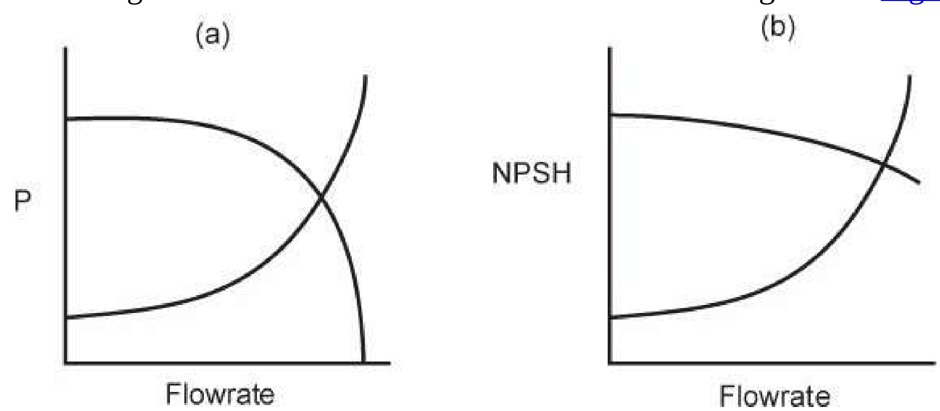

Figure P21.3. Diagrams for Problem 21.3

4. Comment on the following statement: *The film heat transfer coefficient always increases with flowrate to the 0.8 power.*
5. For a given flowrate, it is known that the pump through which the process flows cavitates. The pump has a spare, and a temporary fix to run both pumps simultaneously has been suggested. The pumps can be run either in series or parallel. Which arrangement *might* fix the cavitation problem?
6. A storage tank uses gravity flow to supply a liquid feed material to a holding tank. The pressures and levels in each tank can be considered to be constant. If the viscosity of the fluid is sensitive to temperature (increasing with decreasing temperature), what effect (large or small) will a decrease in temperature have on the flow of the material if the flow is (a) laminar or (b) turbulent? Explain your reasoning.
7. Two columns operate in a chemical plant. The first column operates at a pressure of 10 bar, and the second at a temperature of 1 bar. For process reasons, the pressures in the two columns are both reduced by 0.4 bar. If the internal flows in the columns (mass flows of liquid and vapor) are held constant during this pressure change, which column is more likely to flood?
8. For the separation of a binary mixture in a distillation column, what will be the effect of an increase in column pressure on the following variables?
   a. Tendency to flood at a fixed reflux ratio
   b. Reflux ratio for a given top and bottom purity at a constant number of stages
   c. Number of stages required for a given top and bottom purity at constant reflux ratio
   d. Overhead condenser temperature
9. Explain, using equations where appropriate, why a pump located even with the bottom of a pressurized tank containing a vapor-liquid mixture of a pure component will likely not ever cavitate regardless of the ambient temperature.
10. A storage tank is connected to a pond (at atmospheric pressure!) by a length of 4-in pipe and a gate valve. From previous operating experience, it has been found that when the tank is at a

pressure of 3 atm, the flow through the pipe is 35 m^3/h when the gate valve is fully open. If the pressure in the tank increases to 5 atm, what will be the maximum discharge rate from the tank?

11. Consider the pump and system curves in Figure P21.11 for identical pumps arranged in parallel and series.

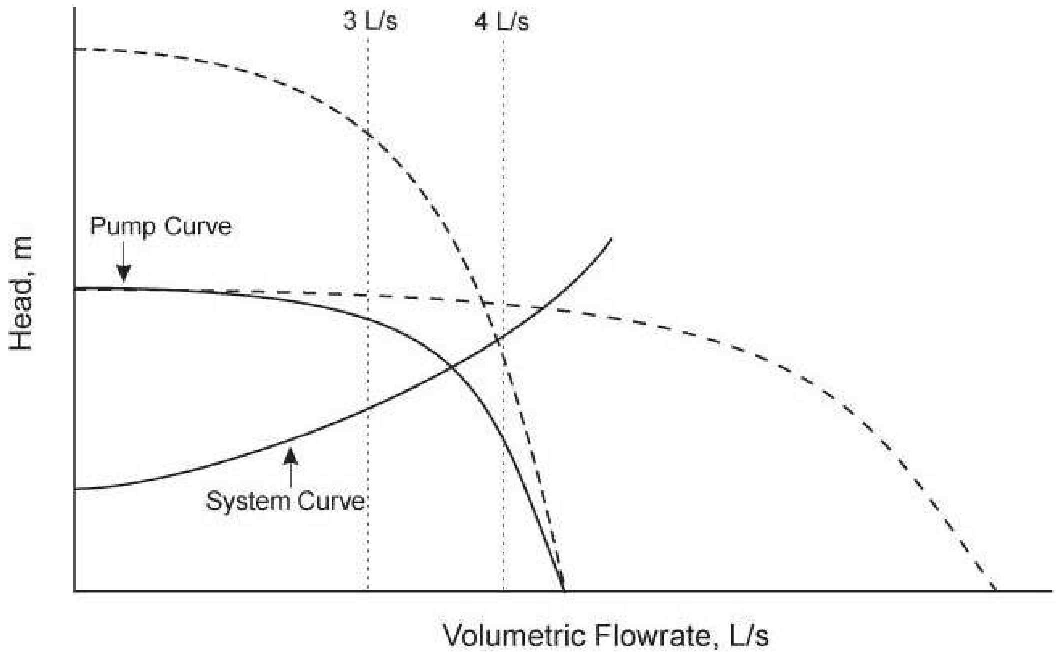

Figure P21.11. Pump and System Curves for Problem 21.11

a. If the flowrate is increased from 3 L/s to 4 L/s, which pump arrangement(s) (single pump, two in series, or two in parallel) will give you the desired flow increase?

b. For your answer(s) to Part (a), will the pump(s) cavitate?

Problems

Background Material for Problems 21.12–21.14

For the steam generator illustrated in Figure 21.1(a) and discussed in Section 21.1, the total resistance to heat transfer, $R_T = 1/(UA)$, is given by the following relationship:

$$R_T = R_i + R_o + R_p + R_{f,i} + R_{f,o}$$

In the above equation,

R_T = total resistance to heat transfer.

R_i = convective resistance of the light oil. This resistance is a function of the flowrate, as shown in Chapter 20. Subscript 1 refers to the base case, and subscript 2 refers to the new case.

$$R_{i2} = R_i(\dot{m}_2/\dot{m}_1)^{0.8}$$

R_o = convective resistance of the boiling water. This resistance is not a function of flowrate. It is a weak function of the temperature drop across the film of vapor on the outside of the tubes. For this analysis, it is assumed constant.

R_p = conductive resistance of the tube walls. This resistance is small and is ignored here.

$R_{f,i}$ and $R_{f,o}$ = fouling resistances on the inside and outside of the tubes, respectively. These two resistances are combined into a single fouling term.

$$R_f = R_{f,i} + R_{f,o}$$

For the conditions given above, the total resistance becomes

$$R_T = R_{i1}(\dot{m}_2/\dot{m}_1)^{0.8} + R_o + R_f$$

The only resistance affected by mass flowrate, $\dot{m}_i$, is R_i.

12. Operating data taken from the plant shortly following the cleaning of the reboiler tubes show that

$$T_{in} = 325°C,\ T_{out} = 293°C,\ \dot{m}_s = 1650\ kg/h,\ T_s = 253°C$$

   a. Find the overall heat transfer coefficient.

   b. Estimate the overall resistance to heat transfer.

13. This is a continuation of Problem 21.12. Estimate the fouling resistance for the base case presented in Figure 21.1(a).

14. This is a continuation of Problem 21.12. Typical heat transfer coefficients given in Table 11.11 are used to predict the ratio of heat transfer resistances between boiling water and light oil. This ratio is estimated to be approximately 1:2. Estimate the individual resistance and heat transfer coefficient for both the oil and the boiling water.

Background Material for Problems 21.15–21.16

The performance curves presented in Figure 21.1 represent those obtained for the heat exchanger used in Problems 21.15–21.16.

15. a. Use the guideline from Tables 11.11 and 11.18 to establish "normal" limits on the amount of heat that could be exchanged.

   b. How do these limits explain the fouling of the heat exchanger observed in the operating unit?

   c. Under the constraints identified in Part (a), what is the highest temperature allowed (flowrates limited to range of *M* values shown)?

16. a. Estimate the relative increase in heat duty that would result from increasing the liquid level from the base case to cover the tubes completely. The input streams and vapor temperature remain the same as in the base case.

   b. Explain why the duty increase is less than the increase in area.

Background Material for Problems 21.17–21.19 is given in Section 21.2

17. For the system presented in Example 21.3 you are asked to appraise the effects of adding a second pump and impeller set identical to the current pump in the system.

   a. Prepare a performance diagram, for the two-pump system and for each impeller size, if the pumps are to be installed in series.

   b. Prepare a performance diagram, for the two-pump system and for each impeller size, if the pumps are to be installed in parallel.

   c. Estimate the pressure achieved for each pump/impeller system in Parts (a) and (b) at flowrates of 400 and 800 gallons/min.

18. Referring to Example 21.3, it has been decided that the storage tank should be kept at 1 bar and the reactor pressure at 1.2 bar. To provide the pressure head required, a pump is placed in line before the heat exchanger. For the 50% increase in flow desired in Example 21.3, answer the following:

   a. What pressure head must be produced by the pump?

b. Which of the pump/impeller systems given in Figure 21.4 is capable of providing sufficient flow?

c. At what location in the system would you place the pump? Explain why.

19. A centrifugal pump produces head by accelerating fluid to a velocity approaching the tip velocity of the rotating impeller. The liquid leaving the pump is at a high velocity and is forced into the discharge pipe from the pump, where it comes into contact with relatively slow-moving fluid in the pipe. As the fluid decelerates, the velocity head is converted into pressure head. For a given pump the following relationships can predict performance:

$$h_2/h_1 = [(N_2/N_1)(D_2/D_1)]^2 \text{ and } \dot{v}_2/\dot{v}_1 = [(N_2/N_1)(D_2/D_1)]^3$$

where

N = speed of rotation of the impeller, rpm

h and $\dot{v}$ = head and volumetric flowrate, respectively

2 and 1 = new and base-case conditions, respectively

D = impeller diameter

The speed of rotation of the impeller can be adjusted, even when a constant speed motor is used, by the use of a pulley system as shown in Figure P21.19.

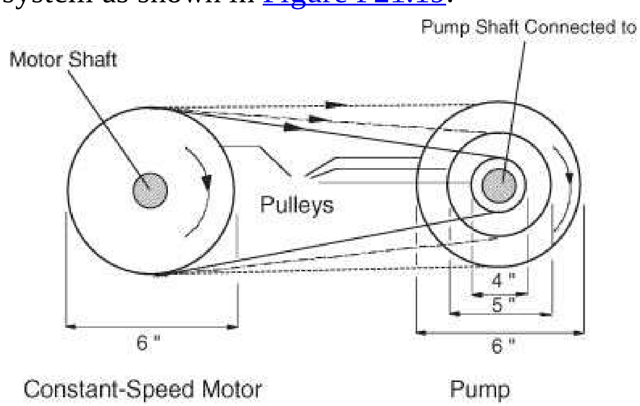

Figure P21.19. Pully System for Problem 21.19

The pump curve given in Figure 21.4 was obtained when the impeller was directly coupled to a constant-speed motor (3500 rpm). Develop performance curves for this pump, with a 7-in-diameter impeller, assuming that the pump is driven via a pulley system in which the constant-speed motor (3500 rpm) has a 6-in-diameter pulley and the motor impeller is connected to a series of pulleys with diameters of 4 in, 5 in, and 6 in. Construct performance diagrams for each pulley combination.

20. Naphthalene is fed to a phthalic anhydride production process. The feed is available at 208°C and 79 kPa. The flowrate is 18,500 kg/h in 1.5-in schedule-40 pipe. A pump with NPSH characteristics as plotted in Figure 21.9 is used. Will this pump be suitable for the desired duty? If not, what modifications would be necessary in order to use the existing pump? Be quantitative.

21. Refer to Example 21.8. There are additional methods other than the two presented for increasing the NPSH in order to avoid pump cavitation. For the following situations, sketch the $NPSH_A$ curve on Figure 21.9, keeping head at 2 m and keeping the flowrate constant. Identify the operating point on the $NPSH_A$ curve in order to determine whether $NPSH_A > NPSH_R$.

a. The diameter of the suction line can be increased. Examine the effect of increasing the diameter of the suction line to 1.25-in schedule 80.

b. The temperature of the toluene can be decreased. Examine the effect of decreasing the temperature of the toluene to 40°C using the original pipe.

22. Refer to the compressor curves given in Figure 21.11.
    a. At a flowrate of 125 Mg/h, what is the exit pressure from the compressor operating at 3500 rpm if the inlet pressure is 2 bar?
    b. If the feed pressure is 4 bar and the compressor is operating at 2200 rpm, how can you obtain an outlet pressure of 9 bar at 125 Mg/h?
    c. It is necessary to raise the pressure from 2 bar to 24 bar at 125 Mg/h. How could this be accomplished at each rpm? Which method do you recommend?
23. A packed scrubber with 20 transfer units has been designed to reduce the solute concentration in 80 kmol/h of air from a mole fraction of 0.01 to 0.006. The solute is absorbed into a 20 kmol/h solvent stream, and it can be assumed that the solvent stream enters solute free. If the flow of solvent is decreased by 15%, what is the new outlet mole fraction of solute in air?
24. Prepare a set of performance curves similar to those in Example 21.10 for a packed scrubber with 10 transfer units that removes solute from air from a mole fraction of 0.02 to a mole fraction of 0.001. The flowrate of the inlet air stream is 80 kmol/h, and the solvent flowrate (assumed solute free) is 40 kmol/h.
25. Repeat Problem 21.24 and Example 21.10 with the following change: In each case, the gas flowrate remains constant at the original conditions. The parameter on the performance curve is now absorber temperature (assumed constant throughout the absorber). There should be five curves for each absorber: the original temperature, an increase of 5°C and 10°C from the original temperature, and a decrease of 5°C and 10°C from the original temperature.
26. The scrubber in Problem 21.23 has been running well for several years. It is now observed that the outlet solute mole fraction in air is 0.002. Suggest at least five reasons for this observation. Suggest at least five ways to compensate for this problem in the short term. Evaluate each compensation method as to its suitability.
27. It is necessary to increase the capacity of an existing distillation column by 25%. As a consequence, the amount of liquid condensed in the condenser must increase by 25%. In this condenser, cooling water is available at 30°C and, under present operating conditions, exits the condenser at 45°C, the maximum allowable return temperature without a financial penalty assessed to your process. Condensation takes place at 75°C. You can assume that the limiting resistance is on the cooling water side. Can the existing condenser handle the scale-up without incurring a financial penalty? What is the new outlet temperature of cooling water? By what factor must the cooling water flow change?
28. In the previous problem, it may be necessary to scale up or scale down by as much as 50% from the original operating conditions. Prepare a performance curve similar to the one in Figure 21.1 for operation of the condenser over this range.

Background Material for Problems 21.29–21.33

In Section 21.3.2, it was shown that Figure 21.19 could be constructed from results of simulating the distillation column. Table P21.29 shows the results of such a simulation for the problem discussed in Section 21.3.2. The data used in Figure 21.19 are contained in this table. The data in Table P21.29 can be used to construct other performance curves, similar to Figure 21.19. In Problems 21.29–21.31, a performance curve should be constructed with the flooding and tray efficiency limits included (given on page 747), as in Figure 21.19.

Table P21.29. Performance Data from a Process Simulator for the Benzene-Toluene Tower

Feed Conc z_F	Benz Rcvry Γ	Feed F kmol/h	Bottm Prod B kmol/h	Reflux Ratio R	Ovhd Vapor V_1 kmol/h	Boil-Up V_{N+1} kmol/h	Liquid Reflux L_0 kmol/h	Cond Duty Q_c GJ/h	Rebl Duty Q_r GJ/h	Bottm Temp °C	Vapor Vel at Top u m/s
0.25	0.80	132.8	102.8	3.30	129.0	122.4	99.0	−3.93	3.87	133.4	1.01
0.30	0.80	110.6	80.6	2.57	107.1	101.7	77.1	−3.26	3.21	132.6	0.84
0.35	0.80	94.8	64.8	2.05	91.5	86.9	61.5	−2.79	2.75	131.8	0.73
0.40	0.80	83.05	53.0	1.66	79.8	75.7	49.8	−2.43	2.39	130.7	0.62
0.25	0.85	124.9	94.9	3.49	134.7	127.8	104.7	−4.10	4.04	134.1	1.05
0.30	0.85	104.1	74.1	2.70	111.0	105.4	81.0	−3.38	3.33	133.5	0.87
0.35	0.85	89.3	59.3	2.14	94.3	89.5	64.3	−2.87	2.83	132.8	0.74
0.40	0.85	78.1	48.1	1.73	81.8	77.6	51.8	−2.49	2.45	132.0	0.64
0.25	0.90	118.0	88.0	3.85	145.4	138.0	115.4	−4.42	4.36	134.9	1.14
0.30	0.90	98.3	68.3	2.96	118.8	112.7	88.8	−3.56	3.61	134.4	0.93
0.35	0.90	84.3	54.3	2.33	99.9	94.8	69.9	−3.04	3.00	133.9	0.78
0.40	0.90	73.8	43.8	1.86	85.5	81.4	55.8	−2.61	2.57	133.3	0.67
0.25	0.95	111.8	81.8	4.78	173.5	164.5	143.5	−5.26	5.20	135.6	1.36
0.30	0.95	93.2	63.2	3.64	139.2	132.1	109.2	−4.23	4.17	135.4	1.09
0.35	0.95	79.9	49.9	2.83	115.0	109.2	85.0	−3.50	3.45	135.1	0.90
0.40	0.95	69.9	39.9	2.24	97.1	92.0	67.1	−2.95	2.91	134.8	0.76

D = 30 kmol/h, x_D = 0.885, P_{bot} = 2 bar, overall tray efficiency = 60%.
Assume ideal system for evaluation of K values and other thermodynamic data.

29. Prepare a performance plot for vapor velocity as a function of feed composition with benzene recovery as a parameter.

30. Prepare a performance plot for boil-up rate as a function of feed composition with benzene recovery as a parameter.

31. Prepare a performance plot for reboiler duty as a function of feed composition with benzene recovery as a parameter.

32. For the distillation problem outlined in Section 21.3.2,

    a. Determine the values of F, B, and x_B required when the concentration of the feed becomes z_F = 0.25. Assume that the values of D and x_D are to remain the same (D = 30 kmol/h and x_D = 0.885) and that the recovery is to remain 90%.

    b. Report any risks associated with the solution.

    c. What are the values of L_0 and V_{N+1} for this case?

33. The information given in Figure 21.18 showed that the benzene recovery column was not operating at maximum benzene recovery.

    a. Estimate the maximum value of benzene recovery for the base case (z_F = 0.321).

b. Give several reasons for operating the current column at less than maximum benzene recovery.

c. Plot a performance curve that shows estimated values for F, L_0, and V_1 at maximum benzene recovery as a function of feed composition.

Background Information for Problems 21.34 and 21.35

In preparing the performance diagrams for the benzene-toluene system in Section 21.3.2, it was noted that under current operation, the feed stream was not introduced at the optimum location. The performance curves, Figure 21.19, were constructed for this nonoptimal feed location.

34. For the same feed, distillate, and bottoms shown in Figure 21.16, determine the following:

a. The boil-up, V_{N+1}, and reflux, L_0, required if the feed is now introduced at the optimum location.

b. Using typical utility costs from Chapter 8, estimate the yearly savings ($/y) obtained by changing the feed location.

35. Assume that the distillation column in Figure 21.16 is to be operated at the same value of L_0, D, and B and the same feed as the base case.

a. Calculate x_D, x_B, and the benzene recovery, Γ, obtained by introducing the feed at the optimum location.

b. Assuming that the benzene lost in the bottom is valued at $0.20/lb, calculate the savings resulting from locating the feed at the optimum location.

36. A heat exchanger was put into service approximately one year ago. The design conditions are that process gas is cooled from 100°C to 50°C, with cooling water entering at 30°C and exiting at 40°C. Initially, the heat exchanger operated satisfactorily, meaning the temperatures for the cooling water and process streams were the same as the design conditions given above. However, over time, it has been observed that, due to impurities in the gas stream, a fouling layer of dirt has built up on the outside of the tubes. This dirt layer has caused the temperatures through the heat exchanger to change, and, in order to maintain the process gas outlet temperature at 50°C with the same design gas flowrate, the mass flow of water has had to be increased to 150% of the design flow, with a corresponding change in cooling water outlet temperature.

Assuming that, for the design case, all the resistance to heat transfer is on the process-gas side, estimate the fouling heat transfer coefficient for the current operation as a fraction or multiple of the gas-phase heat transfer coefficient.

37. In a process that produces a temperature-sensitive product, the final step is to cool the product from 70°C to 35°C prior to sending it to storage. This cooling is achieved using cooling water that is available at 30°C and exiting at 40°C. A shell-and-tube heat exchanger is used with cooling water in the tubes. The process-side resistance is dominant. It is desired to scale down this process. Answer the following questions:

a. If the flowrate of the process fluid decreases to 70% of its design value, by how much must the cooling water flowrate change to maintain the desired exit temperature of the process fluid of 35°C?

b. What is the exit temperature of the cooling water for this case?

38. A desuperheater permits the temperature of saturated steam entering a heat exchanger to be controlled. This is accomplished by having a valve that changes the pressure of the source steam to a known value. When the steam pressure is lowered, the steam becomes superheated. If the

correct amount of bfw is sprayed into the stream, it can be resaturated, and at the lower pressure, the steam is at a lower temperature.

In a particular heat exchanger, condensing steam in the shell at 160°C is used to heat 2 kg/s of a process fluid from 50°C to 100°C. The condensing coefficient is 5000 W/m^2K, and the process-side coefficient is 200 W/m^2K. It is required that the process fluid flowrate be increased by 15%. What steam temperature and flowrate are needed to maintain the outlet process temperature at 100°C?