Chapter 22. Performance of Multiple Unit Operations

What You Will Learn

• How the analysis of the previous chapter can be applied to multiple unit operations

In Chapter 21, techniques for evaluating the performance of individual unit operations were introduced. The use of performance curves was illustrated. Performance curves describe the behavior of a unit over a wide range of inputs. The techniques developed in Chapter 20 were also used to obtain results for a single set of input conditions, which is one point on a performance curve.

In a chemical process, units do not operate in isolation. If the input to one unit is changed, its output changes. However, this output is usually the input to another unit. In certain cases, what is usually considered to be a single unit is actually multiple units operating together. A distillation column is an example of this situation, because the reboiler and condenser are part of the distillation unit. Any change in the column or in either of the heat exchangers will affect the performance of the other two. Therefore, it is necessary to understand the interrelationship between performance of multiple units and to gain experience in analyzing these types of problems.

Given the complexity of the performance of multiple units, performance curves will not be generated. The techniques developed in Chapters 19 through 21 will be used to solve for one point on a performance curve. As in Chapter 21, the case-study approach is used. Four examples of multiple units are illustrated in order to help you understand the approach to analyzing performance problems involving multiple units.

22.1. Analysis of a Reactor with Heat Transfer

Many industrially significant reactions are exothermic. Therefore, in a reactor, it is necessary to remove the heat generated by the heat of reaction. There are several ways to accomplish heat removal. For relatively low heats of reaction, staged adiabatic packed beds (packed with catalyst) with intercooling may be used. Here, no heat is removed in the reaction section of the equipment. Instead, the process fluid is allowed to heat up in the reactor and heat is removed in heat exchangers between short beds of catalyst (reactor sections). For larger exothermic heats of reaction, a shell-and-tube configuration, much like a heat exchanger, is often used. Here, the reaction usually occurs in the tubes, which are packed with catalyst. As the heat of reaction increases, tube diameter is decreased in order to increase the heat transfer area. For extremely large heats of reaction, fluidized beds with internal heat transfer surfaces are often used, due to the constant temperature of the fluid bed and the relative stability of such reactors because of the large thermal mass of well-mixed solid particles.

In all three of these cases, the performance of the reactor and the heat exchanger are coupled. It is not correct to analyze one without the other. In this section, an example of the performance of a shell-and-tube-type reactor (tube bundle containing catalyst immersed in a pool of boiling water) with heat removal is used to illustrate the interrelationship between reactor performance and heat transfer. In Chapter 23, other examples of reactor performance are presented.

A shell-and-tube reactor, as illustrated in Figure 22.1, is used for the following reaction to produce cumene from benzene and propylene.
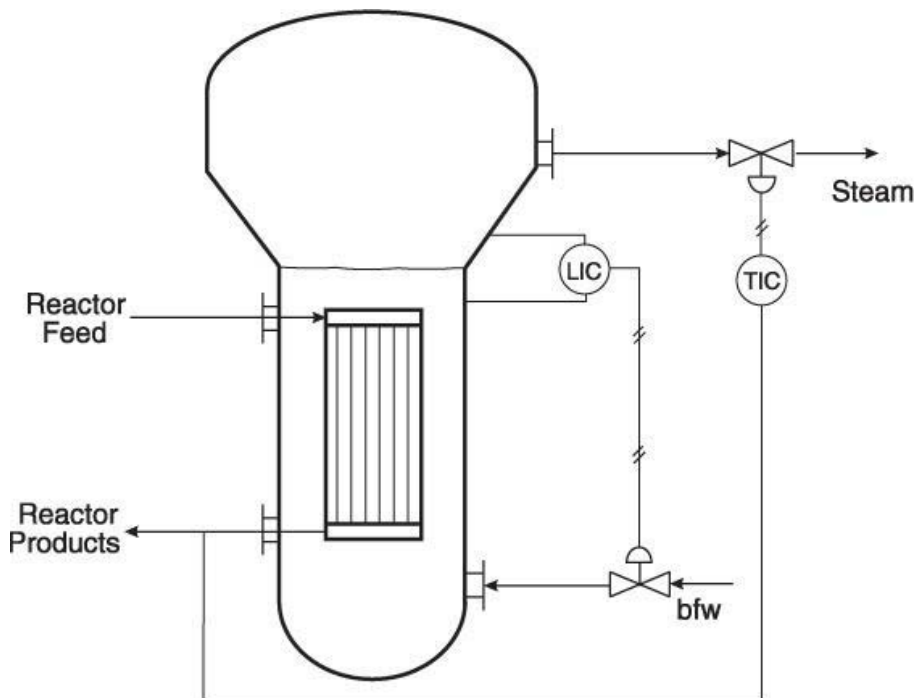
$$C_6H_6 \;+\; C_3H_6 \rightarrow C_9H_{12}$$
$$\text{benzene} \quad \text{propylene} \quad \text{cumene}$$

Figure 22.1. Shell-and-Tube-Type Packed-Bed Reactor

The reaction kinetics are as follows:

$$r = kc_bc_p \quad \text{mol/L s}$$

$$k = 3500 \exp(-13.28 \text{ (kcal/mol)}/RT)$$

Normally, the reaction occurs at 350°C and 3000 kPa (the reaction is not really isothermal; see item 3 below). Heat evolved by the reaction is removed by producing high-pressure steam (4237 kPa, 254°C) from boiler feed water in the reactor shell. Propylene is the limiting reactant, with benzene present in excess. The feed propylene is a raw cut and contains 5 wt% propane impurity. Recently, the propane supplier has been having difficulty meeting specifications, and the propane impurity exceeds 5 wt%. The reduced feed concentration of propylene is causing a decrease in cumene production. In order to maintain the desired cumene production rate, it has been suggested that the reaction temperature be raised to compensate.

The following assumptions are made in order to simplify the problem:

1. There are no other impurities except for the propane.

2. The pressure drop in the reactor is negligible relative to the operating pressure.

3. The temperature profile in the reactor is flat; that is, the temperature is 350°C everywhere in the reactor. This is the most serious assumption, and it is clearly not correct. There will most likely be a hot spot near the reactor entrance. Figure 22.2 illustrates the temperature profiles for the reactor and for the steam. The dotted line is the anticipated temperature profile. By assuming a constant temperature, calculations are simplified, and the concept illustrated in this case study can still be understood.
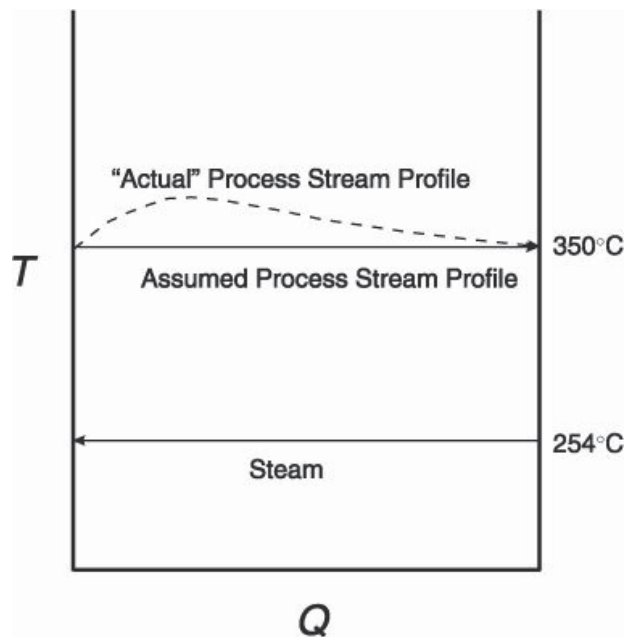
Figure 22.2. Temperature Profiles in Cumene Reactor

Before proceeding with a detailed analysis, it is instructive to understand why a temperature increase was suggested as compensation for decreased propylene feed concentration. If the total feed rate of impure propylene is consistently maintained, the feed rate of propylene decreases with increasing propane impurity. Because propylene is the limiting reactant, cumene production is related to propylene feed by the following relationship:

$$F_c = F_{po} X \qquad (22.1)$$

where

F_c = molar rate of cumene produced

F_{po} = molar rate of propylene feed

X = conversion of propylene

If the molar feed rate of propylene, F_{po}, is reduced, cumene production will decrease unless the conversion, X, is increased. In the reactor, the overall flow of material is constant, because propane replaces propylene; therefore, the space time remains constant. From the reaction expression, a reduced concentration of propylene results in a reduced reaction rate. Intuitively, a reduced reaction rate at constant space time results in reduced conversion. Therefore, from Equation (22.1), the cumene production rate is decreased because of both a decreased reactant feed rate and a decreased conversion.

One method to increase reaction rate is to increase the reactor temperature. The quantitative relationship between temperature increase and cumene production rate can be obtained by the methods learned in a typical reaction engineering class or by use of a process simulator. Example 22.1 illustrates this method.

Example 22.1.

Analyze the heat transfer consequences of increasing the reactor temperature by 10°C and 20°C, increases that correspond to maintaining the desired cumene production rate at approximately 6% and 7% propane impurity, respectively.

The heat transfer performance equation must also be obeyed. Predicting trends shows that the temperature difference between the reaction side and steam side must remain constant:

$$\Delta T(\phi) = \frac{Q(\phi)}{U(\phi)A(\phi)} \qquad \text{(E22.1a)}$$

Because the cumene production rate is to remain constant, the heat load on the reactor remains constant. Because overall flows do not change, the heat transfer coefficient remains constant. Because the reactor is not modified, the heat transfer area remains constant. Therefore, ΔT must remain constant. Because the original ΔT was 96°C, the following table gives the temperatures that are required on the steam side. In order to produce higher-temperature steam, the boiler feed water pressure, and therefore steam pressure, must be increased on the shell side. This is because the temperature at which boiler feed water vaporizes to saturated steam increases with pressure.

Temperature Increase (°C)	Steam Temperature (°C)	Steam Pressure (kPa)
0	254	4237
10	264	5002
20	274	5853

For a 10°C increase in reactor temperature, the pressure on the shell side must increase by almost 20%, and for a 20°C increase in reactor temperature, the pressure on the shell side must increase by about 38%.

Without considering the heat transfer analysis, it would have been easy to suggest the temperature increases be implemented. However, they could not be accomplished without raising the shell-side temperature and pressure. Because the reactor is also a heat exchanger, the performance equation for a heat exchanger must be used in the analysis. Therefore, in order to increase the reactor temperature, the steam temperature, and consequently the steam pressure, must be increased. The unanswered question is whether the reactor can withstand the indicated pressure increases. It is unlikely that a vessel designed for a high pressure like 4237 kPa would be overdesigned to handle a 20% increase in pressure, and a design for 38% increase would be extremely uncommon.

It is instructive to continue this analysis and determine whether there are other ways to increase the conversion in the reactor to compensate for the reduced feed, and this is illustrated in Example 22.2.

Example 22.2.

Suggest another method for increasing the conversion in the reactor so that the cumene production rate remains constant with reduced propylene feed concentration.

The question to be answered is what can be adjusted, other than temperature, that will affect the reaction rate and/or the conversion. In the reaction rate expression, the primary temperature effect is in the Arrhenius expression. For gas-phase reactions, assuming ideal gas behavior, the concentrations $c_i = P_i/RT$. The increase in temperature discussed in Example 22.1 will actually reduce the concentrations; however, the exponential increase in the reaction rate constant is the primary effect. It is also observed that increasing the pressure increases the concentration, thereby increasing the reaction rate. There is an additional effect. This can be seen from the equation of a plug flow reactor.

$$\frac{\tau}{c_{Ao}} = \int_0^X \frac{dX}{-r_A}$$

Because the mass flowrate $\dot{m} = \rho A v$, and increasing the pressure increases the density, at constant mass flowrate, the velocity drops. Therefore, the residence time, τ, in the reactor increases, which results in increased cumene conversion (upper limit of integral increases) and increased cumene production. Of course, pressure limitations of the equipment in the process or in the tube side of the reactor may limit the pressure increase possible.

In this case study, the most likely solution to the problem is some combination of temperature and pressure increase. If this is insufficient, it may be necessary to try to increase the feed rates of the reactants. However, consistent with the lesson of this case study, the performance of the pumping equipment may limit possible feed rate increases.

In summary, this case study introduces the concept that performance of any unit cannot be considered in isolation. For the reactor, it seems straightforward to do an analysis involving only reaction kinetics and conclude that a temperature increase is a solution to the problem of reduced feed. However, it is now clear that because the reactor is also a heat exchanger, both reactor and heat exchanger analyses must be done simultaneously in order to obtain a correct result. In order to get the correct numerical result, the correct temperature profile on the reaction side must be used. It is obtained by using a process simulator or by solving the differential equations for conversion and temperature simultaneously. The correct temperature profile, as illustrated in Figure 22.2, arises because of the reaction rate, $r \propto e^{-E/RT}$. Therefore, as the temperature increases, r increases by a large amount, generating even more heat, and the peak in the temperature profile may increase significantly, causing hot spots to develop. Hot spots can damage catalyst or promote undesired side reactions. Chapter 23 treats these issues in more detail.

22.2. Performance of a Distillation Column

A distillation column requires a reboiler to add the energy necessary to accomplish the separation. A condenser is also required to reject heat to the surroundings. The performance of a distillation column cannot be analyzed without consideration of the performance of both heat exchangers. This is illustrated in the following case study.

The distillation column illustrated in Figure 22.3 is used to separate benzene and toluene. It contains 35 sieve trays, with the feed on Tray 18. The relevant flows are given in Table 22.1. Your assignment is to recommend changes in the tower operation to handle a 50% reduction in feed. Overhead composition must be maintained at 0.996 mole fraction benzene. Cooling water is used in the condenser, entering at 30°C and exiting at 45°C. Medium-pressure steam (185°C, 1135 kPa) is used in the reboiler.
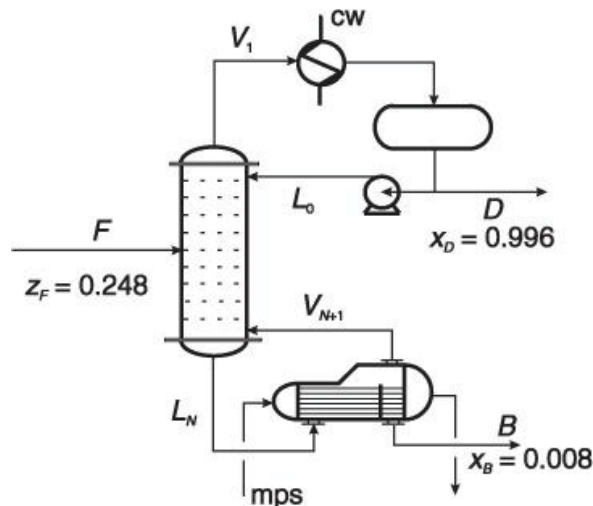
Figure 22.3. Distillation of Benzene from Toluene

Table 22.1. Operating Conditions for Process in Figure 22.3

Input/Output	Flow (kmol/h)	Mole Fraction Benzene	Temperature (°C)
Inputs			
Feed, F	141.3	0.248	90
Reflux, L_0	130.7	0.996	112.7
Boil-up, V_{N+1}	189.5	0.008	145.3
Outputs			
Distillate, D	34.3	0.996	112.7
Still bottoms, L_N	296.5	0.008	145.3
Bottoms product, B	107.0	0.008	145.3

The operating conditions of the tower before reduction in feed are in Table 22.1.

Among the several possible operating strategies for accomplishing the necessary scale-down are the following:

1. Scale down all flows by 50%. This is possible only if the original operation is not near the lower velocity limit that initiates weeping or reduced tray efficiency. If 50% reduction is possible without weeping, reduced tray efficiency, or poor heat-exchanger performance, this is an attractive option.

2. Operate at the same boil-up rate. This is necessary if weeping or reduced tray efficiency is a problem. The reflux ratio must be increased in order to maintain the reflux necessary to maintain the same liquid and vapor flows in the column. In this case, weeping, lower tray efficiency, or poor heat-exchanger performance caused by reduced internal flows is not a problem. The downside of this alternative is that a purer product will be produced and unnecessary utilities will be used.

In Example 22.3, the analysis will be done by assuming that it is possible to scale down all flows by 50% without weeping or reduced tray efficiency.

Example 22.3.

Estimate the pressure drop through the column. To what weir height does this correspond?

Interpolation of tabulated data [1] yields the following relationships for the vapor pressures in the temperature range of interest. Note that the temperatures predicted by these interpolated relationships differ slightly from those predicted by the simulator results in Table 22.1.

$$\ln P^* \ (kPa) = 15.1492 - \frac{3706.84}{T(K)} \quad \text{benzene}$$

$$\ln P^* \ (kPa) = 15.3877 - \frac{4131.14}{T(K)} \quad \text{toluene}$$

It is assumed that the bottom is pure toluene and the distillate is pure benzene. This is a good assumption for estimating top and bottom pressures given the mole fractions specified for the distillate and bottoms. Therefore, at the bottom temperature of 141.7°C, the vapor pressure of toluene, and hence the pressure at the bottom of the column, is 227.2 kPa. At the top temperature of 104.2°C, the vapor pressure of benzene, and hence the pressure at the top of the column, is 204.8 kPa.

Because there are 35 trays, the pressure drop per tray is

$$\Delta P = (227.2 - 204.8)/35 = 0.64 \ \text{kPa/tray}$$

If it assumed that the weir height is the major contribution to the pressure drop on a tray, then

$$\Delta P = \rho g h$$

where h is the weir height. Assuming an average density of 800 kg/m³, then

$$h = (640 \ \text{Pa})/[(800 \ \text{kg/m}^3)(9.8 \ \text{m/s})] = 0.08 \ \text{m} \approx 3 \ \text{in}$$

This is a typical weir height and is consistent with the assumption that the weir height is dominant.

The pressure drop is assumed to remain constant after the scale-down because the weir height is not changed. In practice, there is an additional contribution to the pressure drop due to gas flow through the tray orifices. This would change if column flows changed, but the pressure drop through the orifices is small and should be a minor effect. Examples 22.4 and 22.5 illustrate how the performance of the reboiler and condenser affect the performance of the distillation column.

Example 22.4.

Analyze the reboiler to determine how its performance is altered at 50% scale-down.

Figure E22.4 shows the T-Q diagram for this situation. Because the amounts of heat transferred for the base and new cases are different, the Q values must be normalized by the total heat transferred in order for these profiles to be plotted on the same scale. The solid lines are the original case (subscript 1). For the new, scaled-down case (subscript 2), a ratio of the energy balance on the reboiled stream for the two cases yields

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_2 \lambda}{\dot{m}_1 \lambda} \tag{E22.4a}$$

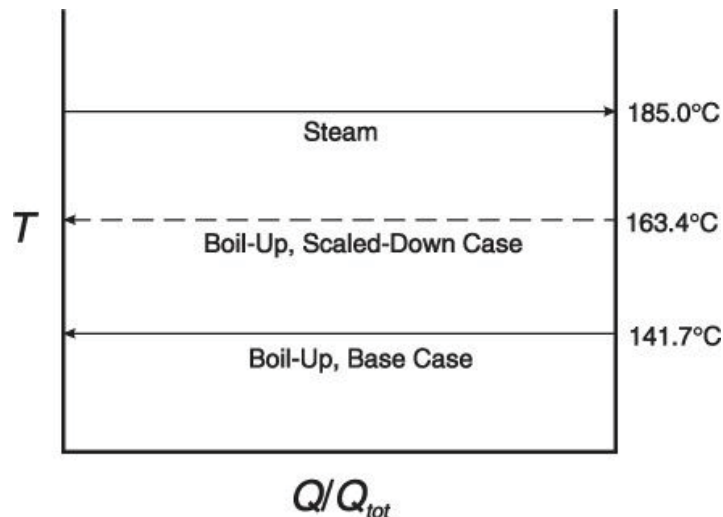**Figure E22.4. Temperature Profiles in Reboiler for Example 22.4**

If it is assumed that the latent heat is unchanged for small temperature changes, then $Q_2/Q_1 = 0.5$, because at 50% scale-down, the ratio of the mass flowrates in the reboiler is 0.5. The ratio of the heat transfer equations yields

$$\frac{Q_2}{Q_1} = 0.5 = \frac{U_2 A \Delta T_2}{U_1 A \Delta T_1} \approx \frac{\Delta T_2}{43.3} \qquad \text{(E22.4b)}$$

In Equation (E22.4b), it is assumed that the overall heat transfer coefficient is constant. It is assumed here that $U \neq f(T)$. This assumption should be checked, because for boiling heat transfer coefficients with large temperature differences, the boiling heat transfer coefficient may be a strong function of temperature difference.

From Equation (E22.4b), it is seen that $\Delta T_2 = 21.7°C$. Therefore, for the reboiler to operate at 50% scale-down, with the steam side maintained constant, the boil-up temperature must be 163.4°C. This is shown as the dotted line in Figure E22.4.

From the vapor pressure expression for toluene, the pressure at the bottom of the column is now 372.5 kPa. It would have to be determined whether the existing column could withstand this greatly increased pressure.

Example 22.4 shows that the reboiler operation requires that the pressure of the boil-up stream be increased. This is not the only possible alternative. All that is necessary is that the temperature difference be 21.7°C. This could be accomplished by reducing the temperature of the steam to 163.4°C. In most chemical plants, steam is available at discrete pressures, and the steam used here is typical of medium-pressure steam. Low-pressure steam would be at too low a temperature to work in the scaled-down column. However, the pressure and temperature of medium-pressure steam could be reduced if there were a throttling valve in the steam feed line to reduce the steam pressure. The resulting steam would be superheated, so desuperheating would also be necessary. This could be accomplished by spraying water into the superheated steam. A change in a utility stream is almost always preferred to a change in a process stream. The condenser is now analyzed in Example 22.5.

Example 22.5.

Using the results of Example 22.4, analyze the condenser for the scaled-down case.

Under the assumption that the pressure drop for the column remains at 22.4 kPa, the pressure in the

condenser is now 350.0 kPa. From the vapor pressure expression for benzene, the new temperature in the condenser is 126.0°C. The remainder of the analysis is similar to Example 21.1. The *T-Q* diagram is illustrated in Figure E22.5. Because an organic is condensing, it is assumed that the resistance on the cooling water side is approximately equal to the resistance on the condensing side. Therefore,
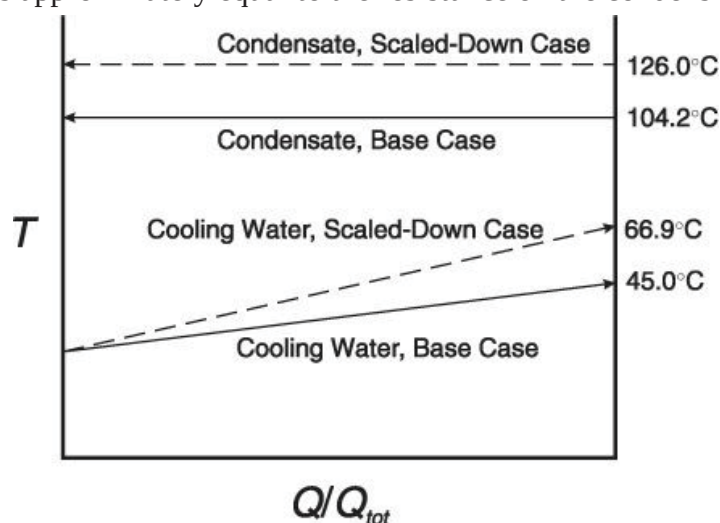

Figure E22.5. Temperature Profiles in Condenser for Example 22.5

$$\frac{U_2}{U_1} = \frac{\dfrac{1}{h_{o1}} + \dfrac{1}{h_{i1}}}{\dfrac{1}{h_{o2}} + \dfrac{1}{h_{i2}}} = \frac{\dfrac{2}{h_{o1}}}{\dfrac{1}{h_{o1}} + \dfrac{1}{h_{o1}M^{0.8}}} = \frac{2}{1 + \dfrac{1}{M^{0.8}}} \qquad \text{(E22.5a)}$$

where $M = \dot{m}_2/\dot{m}_1$.

The ratio of the base cases for the energy balance on the condensing stream is

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_2 \lambda}{\dot{m}_1 \lambda} \qquad \text{(E22.5b)}$$

Because the ratio of the mass flowrates is 0.5, and assuming that the latent heat is unchanged with temperature in the range of interest, $Q_2/Q_1 = 0.5$.

The ratio of the base cases for the energy balance and the heat-exchanger performance equation are, respectively,

$$0.5 = M\frac{T - 30}{15} \qquad \text{(E22.5c)}$$

$$0.5 = \frac{2(T - 30)}{(66.42)\left(1 + \dfrac{1}{M^{0.8}}\right)\ln\left(\dfrac{96}{126 - T}\right)} \qquad \text{(E22.5d)}$$

Solution of Equations (E22.5c) and (E22.5d) yields

$$M = 0.203$$

$$T = 66.9°C$$

Therefore, the cooling water rate must be reduced to 20% of the original rate, and the outlet water temperature is increased by about 22°C. This would cause increased fouling problems on the cooling water side. Therefore, it may be better to use a higher reflux ratio rather than reduce the flows in the column.

Observation indicates that in order to operate the distillation column at 50% scale-down with process flows reduced by 50%, a higher pressure is required and the cooling water will be returned at a significantly higher temperature than before scale-down. The operating pressure of the column is determined by the performance of the reboiler and condenser.

Examples 22.4 and 22.5 reveal a process bottleneck at the reboiler that must be resolved in order to accomplish the scale-down. In the process of changing operating conditions (capacity) in a plant, a point will be reached where the changes cannot be increased or reduced any further. This is called a **bottleneck**. A bottleneck usually results when a piece of equipment (usually a single piece of equipment) cannot handle additional change. In this problem, one bottleneck is the high cooling water return temperature, which would almost certainly cause excessive fouling in a short period of time. Another potential bottleneck is tray weeping due to the greatly reduced vapor velocity. In addition to the solution presented here, there are a variety of other possible adjustments, or debottlenecking strategies, which follow with a short explanation for each. Another debottlenecking problem is presented in Chapter 24.

1. **Replace the Heat Exchangers:** Equation (E22.4b) shows that a new heat exchanger with half of the original area allows operation at the original temperature and pressure. The heat transfer area of the existing exchanger could be reduced by plugging some of the tubes, but this modification would require a process shutdown. This involves both equipment downtime and capital expense for a new exchanger. Your intuitive sense should question the need to get a new heat exchanger to process less material. You can be assured that your supervisor would question such a recommendation.

2. **Keep the Boil-Up Rate Constant:** This would maintain the same vapor velocity in the tower. If the operation occurs near the lower velocity limit that initiates weeping or lower tray efficiency, this is an attractive option. The constant boil-up increases the tower separation. This is an attractive option that results in much smaller temperature and pressure changes.

3. **Introduce Feed on a Different Plate:** This must be combined with Option 4. The plate should be selected to decrease the separation and increase the bottoms concentration of the lower boiling fraction. This lowers the process temperature, increases the ΔT for heat transfer and the reboiler duty. This may or may not be simple to accomplish depending on whether the tower is piped to have alternate feed plates. As in Option 2, the reflux and coolant inputs will change.

4. **Recycle Bottoms Stream and Mix with Feed:** This must be combined with Option 3. By lowering the concentration of the feed, the concentration of the low boiler in the bottoms can be increased, the temperature lowered, and the reboiler duty increased. This represents a modification of process configuration and introduces a new (recycle) input stream into the process.

   The adjustments outlined were based on the following:

   1. Replace the exchanger (Option 1).

   2. Modify utility stream (discussed after Example 22.4).

   3. Modify process stream (Option 2).

   4. Modify equipment (Option 3 or Option 1, if plugging of tubes to reduce area is chosen).

   5. Modify process configuration (Option 4).

   6. Modify operating conditions T and P (as in Examples 22.4 and 22.5).

It has been shown that there are a number of paths that remove the bottleneck that exists at the reboiler. Adjustments 1 and 2 required no changes in process inputs. All others required modifications

of several input streams to maintain the output quality.

It should be noted that, at 50% scale-down, it is possible that weeping or low tray efficiency may be observed (see Section 21.3.2 and Figure 21.19). This must be considered before recommending such a scale-down. Furthermore, reduction of the cooling water flowrate in the condenser, combined with increased cooling water temperatures, could cause fouling problems, as pointed out in Example 22.5. Care must taken to understand the consequences of process modifications.

During plant operations, when seeking the best operating conditions, incremental changes can be made and the effects observed. This is termed "evolutionary operation" [2]. Amoco has reported a significant increase in its annual income from evolutionary operation [3]. In the design case, changes recommended are necessarily conservative to ensure that equipment purchased and installed is adequate to meet the conditions when the plant is operated.

22.3. Performance of a Heating Loop

In Section 22.2, it was noted how the performance of a distillation column was affected by the performance of the reboiler and condenser. It was possible to analyze the reboiler, column, and condenser sequentially in order to solve the scale-down problem. In this section, the performance of a reactor in which the large, exothermic heat of reaction is removed by a heat transfer fluid is examined. It will be seen that the bottleneck for scale-up of this process is the performance of the heat removal loop. The analysis of this problem is more complex than for the distillation column, because all units involved must be analyzed simultaneously in order to solve the problem.

The problem to be analyzed is illustrated in Figure 22.4. It is a part of the allyl chloride problem discussed in more detail in Appendix C, on the accompanying CD.
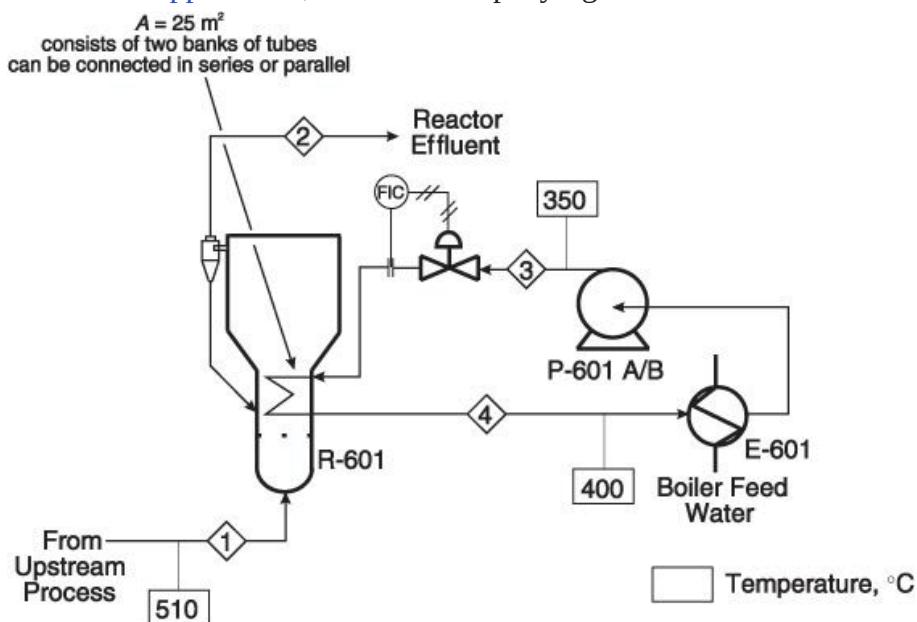

Figure 22.4. Dowtherm A Loop for Allyl Chloride Reactor

Due to an emergency, an unscheduled shutown at an another allyl chloride production plant operated by your company, it is necessary to increase production temporarily at your plant. Your job is to determine the maximum scale-up possible for the allyl chloride production reactor. The reactor is a fluidized bed, with a cyclone for solids recovery, operating isothermally at 510°C. The reactor is operating at two times the minimum fluidization velocity. Based on the cyclone design and the solids

handling system, it is known that an increase of at least 100% of process gas flow can be accommodated. The reaction proceeds to completion as long as the reactor temperature remains at 510°C. A temperature increase destroys the catalyst, and a temperature decrease results in incomplete reaction and undesired side products. The reaction is exothermic and heat is removed by Dowtherm A circulating in the loop. The maximum operating temperature for Dowtherm A is 400°C, and its minimum operating pressure at this temperature is 138 psig. There are two heat-exchanger units in the reactor, and they are currently operating in series. The heat transfer resistance on the reactor side is four times that on the Dowtherm A side. The inside of the reactor is always at 510°C due to the well-mixed nature of the fluidized solids. There is a spare pump. High-pressure steam is made from boiler feed water in the heat exchanger. The boiler feed water is available at 90°C and has been pumped to 600 psig by a pump that is not shown. The pool of liquid being vaporized can be assumed to be at the vaporization temperature, 254°C. All resistance in the heat exchanger is assumed to be on the Dowtherm A side. At normal operating conditions, Dowtherm A circulates at 85 gal/min in the loop. The temperatures of each stream are shown on Figure 22.4. The pump curve for the pump in the loop is shown on Figure 22.5.
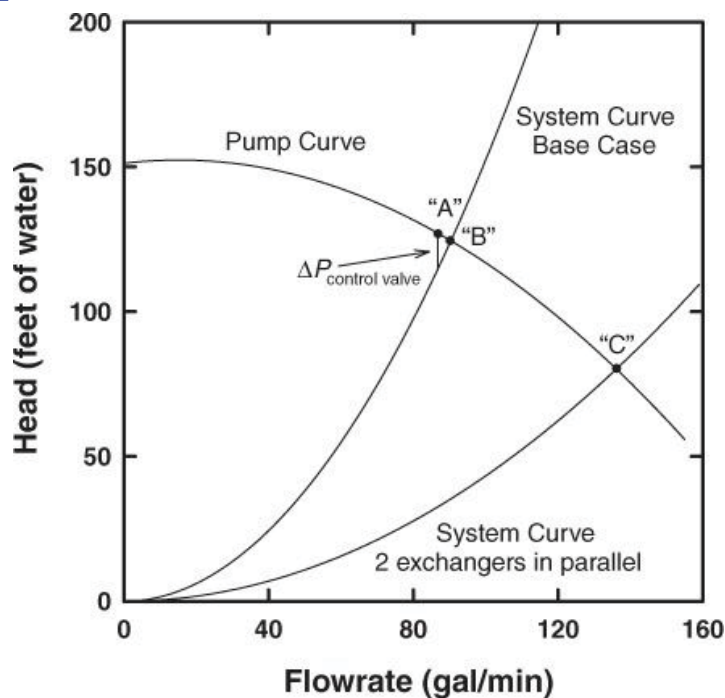

Figure 22.5. Performance of Dowtherm A Loop

An energy balance on the process fluid in the reactor yields

$$Q_R = \dot{n}_{Cl}\Delta H_{rxn} \tag{22.2}$$

where $\dot{n}_{Cl}$ is the molar flowrate of chlorine (the limiting reactant), and the subscript R refers to the reactor. The heat of reaction term in Equation (22.2) is on a unit mole basis. Because the heat of reaction is constant during scale-up, Equation (22.2) shows that the amount of heat removed increases proportionally with the amount of scale-up. From this energy balance, there is nothing to suggest that the reactor cannot be scaled up by a factor of 2, the maximum allowed by the reactor. However, the question that must be answered is whether twice the heat can be removed from the reactor by the heat-exchange loop. An energy balance on the Dowtherm A in the reactor yields

$$Q_R = \dot{m}_D C_{pD}(T_4 - T_3) \tag{22.3}$$

where the subscript D refers to the Dowtherm A. From Equation ([22.3](#)), it is observed that the heat removal rate is proportional to both the mass flowrate and the temperature increase of the Dowtherm A. Therefore, the most obvious method is to increase the flowrate of Dowtherm A, so it is necessary to determine the maximum flowrate possible for Dowtherm A. [Example 22.6](#) illustrates this calculation.

Example 22.6.

For the pump curve provided in [Figure 22.5](#), determine the maximum possible flowrate for Dowtherm A in the loop. The pump operates at only one speed, described by the curve given in the figure. The pressure drops between streams are shown in [Figure E22.6](#).
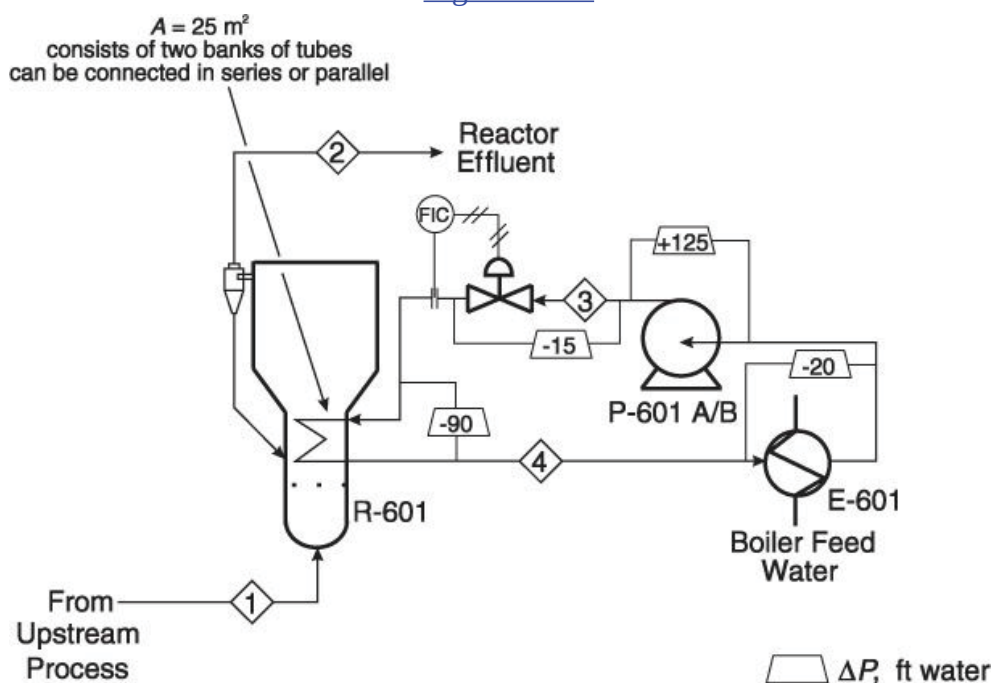
Figure E22.6. Pressure Drops in Dowtherm A Loop for Allyl Chloride Reactor

The pump operates only at conditions shown on the pump curve. The pressure drop in the system and regulating valves must equal the pressure delivered by the pump. The base-case solution gives one point on this line (point "a"). Of this total pressure drop, 15 feet of water is from the control valve, which can be adjusted independently. The remaining pressure drop, 110 feet of water (125 − 15 feet of water), is the **system pressure drop** and is dependent on the flowrate through the system. The system pressure drop can be obtained from the equation for frictional pressure losses.

$$\Delta P = \frac{2\rho f L_{eq} v^2}{D} \qquad \text{(E22.6a)}$$

Taking the ratio of Equation (22.6a) for the scaled-up case (subscript 2) to the base case (subscript 1) for constant L, D, and the assumed constant friction factor (high Reynolds number, fully developed turbulent flow) yields

$$\frac{\Delta P_2}{110} = \frac{v_2^2}{v_1^2} = \frac{\dot{m}_2^2}{\dot{m}_1^2} \qquad \text{(E22.6b)}$$

where M is the scale-up factor, the ratio of the scaled-up Dowtherm A flowrate to that for the base case. Equation (22.6b) is plotted with the pump curve on [Figure 22.5](#). The maximum flowrate occurs when the valve is wide open. When the valve is wide open, there is no (or very little) pressure drop across the valve. This is shown as point "b" and occurs at 89 gal/min.

shows that the maximum scale-up is limited to about 5% by the Dowtherm A flowrate in the loop. A bottleneck has been identified. In order for additional scale-up to be possible, a method for increasing the maximum Dowtherm A flowrate must be found. Several options are possible:

1. Run only one pump, but operate the reactor heat exchangers in parallel.
2. Run both pumps (the operating pump and the spare) in parallel, with the reactor heat exchangers in series.
3. Run both pumps in series, with the reactor heat exchangers in series.
4. Run both pumps (series or parallel), and operate the reactor heat exchangers in parallel.

Option 1 is illustrated in . The remaining options are the subject of a problem at the end of the chapter.

Example 22.7.

Determine the maximum possible Dowtherm A flowrate for one pump operating with the two reactor heat exchangers operating in parallel.

From Equation (E22.6a), it is seen that

$$\Delta P \propto L_{eq} v^2 \qquad \text{(E22.7a)}$$

The equivalent length in the heat exchanger is reduced by half for operation in parallel. Half the flowrate goes through each heat exchanger, which means that the Dowtherm A has half the original velocity. Therefore, the pressure drop through the reactor heat exchangers drops by a factor of 8 and becomes 11.25 feet of water. The base-case pressure drop is now 31.25 feet of water. Therefore, a new system curve is obtained. Its equation is

$$\Delta P = 31.25\, M^2 \qquad \text{(E22.7b)}$$

The new system curve along with the pump curve is plotted on Figure 22.5. Point "c" represents the maximum possible flowrate of Dowtherm A, 135 gal/min. This represents a 58.8% scale-up.

Therefore, if only the original pump is used with the reactor heat exchangers in parallel, the Dowtherm A circulation rate can be increased by almost 50%. For Options 2 through 4, additional increases are possible. It should be noted that liquid velocities exceeding 10 ft/sec are not recommended due to potential erosion problems. Therefore, if the original case was designed for the maximum recommended velocity, operation at a 50% higher velocity for an extended period of time is not a good idea. Proceeding with this problem, the assumption is that the original design permits 50% scale-up without exceeding the maximum recommended velocity.

Because the flowrate of the Dowtherm A can be scaled up by 50% does not mean that the heat removal rate can be scaled up by the same factor. Analysis of the heat transfer in both heat exchangers is required. For the reactor, the energy balances were given by Equations (22.2) and (22.3), for the process side and the Dowtherm A side, respectively. The performance equation for the reactor heat exchanger is

$$Q_R = U_R A_R \Delta T_{lm} \qquad \text{(22.4)}$$

For the heat exchanger producing steam, there are two additional equations. One is the energy balance on the boiler feed water to steam side.

$$Q_h = \dot{m}_s (C_{p,bfw} \Delta T + \lambda) \qquad \text{(22.5)}$$

where $\dot{m}_S$ is the mass flowrate of steam, ΔT is the temperature difference between the boiler feed water inlet and the vaporization temperature, and λ is the latent heat of vaporization of the steam. The

second equation is the performance equation for the heat exchanger.

$$Q_h = U_h A_h \Delta T_{lm} \tag{22.6}$$

The energy balance on the Dowtherm A side is identical (with the temperature difference reversed) to Equation (22.3). It is assumed that all heat removed from the reactor is transferred to make steam ($Q_R = Q_h$).

Therefore, there are five independent equations that describe the performance of the loop. The unknowns are T_3, T_4, $Q_R = Q_h$, $\dot{m}_p$, $\dot{m}_D$, and $\dot{m}_s$. For a given value of $\dot{m}_D$, the mass flowrate of Dowtherm A, a unique solution exists. The method for obtaining this solution is illustrated in Example 22.8.

Example 22.8.

From a heat transfer analysis, determine the maximum scale-up possible for the allyl chloride reactor for the original case and for the case of one pump with the reactor heat exchangers operating in parallel.

Ratios of the scaled-up case (subscript 2) to the base case (subscript 1) will be used. Q_2/Q_1 is the ratio of the heat transfer rates, M is the ratio of the mass flowrates of Dowtherm A, and M_s is the ratio of the mass flowrates of steam. Only three of the five equations are coupled. Base-case ratios of Equations (22.3), (22.4), and (22.6) must be solved for T_3, T_4, and Q_2/Q_1. Then M_s and M are determined from Equations (22.5) and (22.2), respectively. The value of Q_2/Q_1 defines the level of scale-up possible.

The base-case ratios for Equations (22.3), (22.4), and (22.6) are

$$\frac{Q_2}{Q_1} = \frac{M(T_4 - T_3)}{50} \tag{E22.8a}$$

$$\frac{Q_2}{Q_1} = \frac{5(T_4 - T_3)}{133.44\left(4 + \dfrac{1}{M^{0.8}}\right)\ln\left(\dfrac{510 - T_3}{510 - T_4}\right)} \tag{E22.8b}$$

$$\frac{Q_2}{Q_1} = \frac{M^{0.8}(T_4 - T_3)}{119.26\ln\left(\dfrac{T_4 - 254}{T_3 - 254}\right)} \tag{E22.8c}$$

For the case of maximum scale-up with the original heat-exchanger configuration, $M = 89/85 = 1.05$. The solutions are

$$T_3 = 348.7°C$$

$$T_4 = 397.4°C$$

$$Q_2/Q_1 = 1.02$$

From Equation (22.2), because $Q_2/Q_1 = 1.02$, and because the heat of reaction is constant, the process flowrate increases by 2%. So only 2% scale-up is possible.

For one pump with the reactor heat exchangers in parallel, the mass flowrate in each reactor heat exchanger is half of the total flow. Therefore, Equation (E22.8b) becomes

$$\frac{Q_2}{Q_1} = \frac{5(T_4 - T_3)}{133.44\left(4 + \left(\dfrac{2}{M}\right)^{0.8}\right)\ln\left(\dfrac{510 - T_3}{510 - T_4}\right)} \tag{E22.8d}$$

For $M = 1.588$, solution of Equations (E22.8a), (E22.8c), and (E22.8d) yields

$$T_3 = 331.7°C$$
$$T_4 = 367.8°C$$
$$Q_2/Q_1 = 1.15$$

Therefore, 15% scale-up is possible.

Performance of the heat-exchange loop is illustrated on the *T-Q* diagram shown in Figure 22.6. The lines for the reactor and the steam-generating heat exchanger are unchanged. The line for the Dowtherm A changes slope as the mass flowrate, the reactor heat-exchanger configuration, or the number and configuration of pumps change.
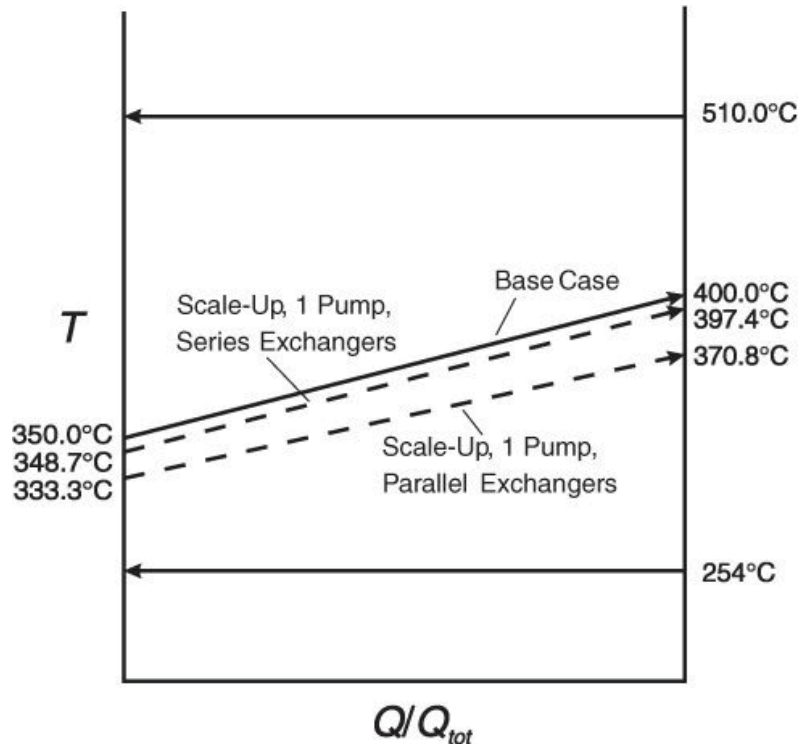

Figure 22.6. Temperature Profiles for Exchangers in Dowtherm A Loop

This complex problem illustrates an important feature of chemical processes. Often the bottleneck to solving a problem is elsewhere in the process. In this problem, the bottleneck to reactor scale-up is not in the reactor itself, but in the heat removal loop. Nevertheless, several alternatives are available to increase the Dowtherm A flowrate and scale up the process.

22.4. Performance of the Feed Section to a Process

A very common feature of chemical processes is the mixing of reactant feeds prior to entering a reactor. When two streams mix, they are at the same pressure. The consequences of this are illustrated by the following scenario.

Phthalic anhydride can be produced by reacting naphthalene and oxygen. The feed section to a phthalic anhydride process is shown in Figure 22.7. The mixed feed enters a fluidized bed reactor operating at five times the minimum fluidization velocity. A stream table is given in Table 22.2. It is assumed that all frictional pressure losses are associated with equipment and that frictional losses in the piping are negligible. It is temporarily necessary to scale down production by 50%. Your job is to

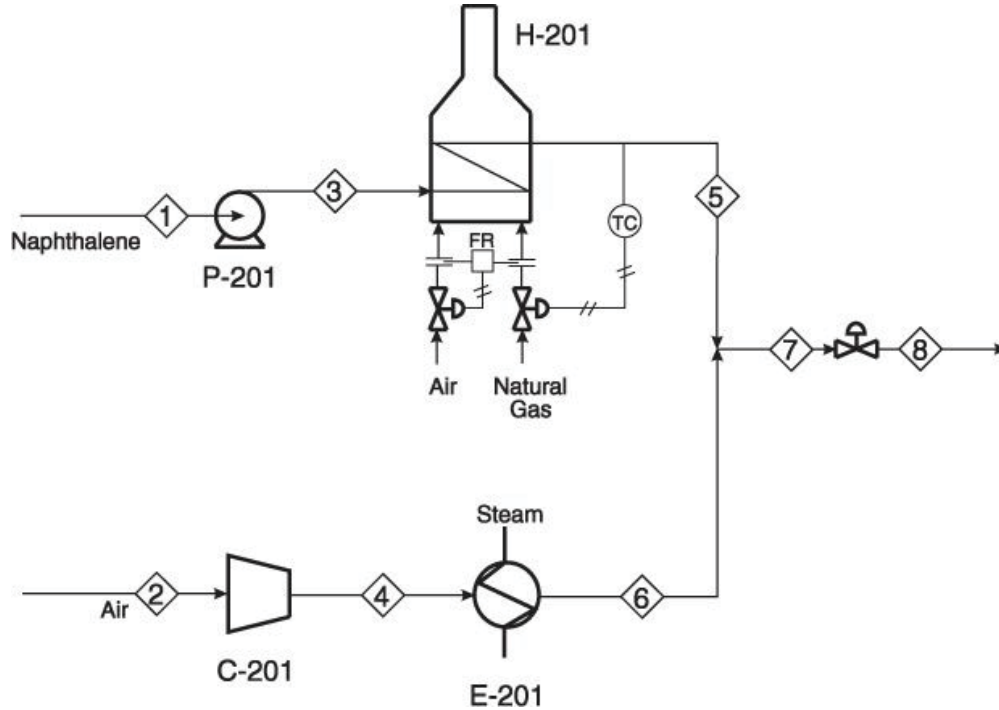determine how to scale down the process and to determine the new flows and pressures.

Figure 22.7. Feed Section to Phthalic Anhydride Process

Table 22.2. Partial Stream Table for Feed Section in <u>Figure 22.7</u>

	Stream							
	1	2	3	4	5	6	7	8
P (kPa)	80.00	101.33	343.00	268.00	243.00	243.00	243.00	200.00
Phase	L	V	L	V	V	V	V	V
Naphthalene (Mg/h)	12.82	—	12.82	—	12.82	—	12.82	12.82
Air (Mg/h)	—	151.47	—	151.47	—	151.47	151.47	151.47

It is necessary to have pump and compressor curves in order to do the required calculations. In this example, equations for the pump curves are used. These equations can be obtained by fitting a polynomial to the curves provided by pump manufacturers. As discussed in <u>Chapter 21</u>, pump curves are usually expressed as pressure head versus volumetric flowrate. This is so that they can be used for a liquid of any density. In this section, pressure head and volumetric flowrate have been converted to absolute pressure and mass flowrate using the density of the fluids involved. Pump P-201 operates at only one speed, and an equation for the pump curve is

$$\Delta P(\text{kPa}) = 500 + 4.663\dot{m} - 1.805\dot{m}^2 \quad \dot{m} \leqslant 16.00 \text{ Mg/h} \tag{22.7}$$

Compressor C-201 operates at only one speed, and the equation for the compressor curve is

$$\frac{P_{out}}{P_{in}} = 5.201 + 2.662 \times 10^{-3}\dot{m} - 1.358 \times 10^{-4}\dot{m}^2 \tag{22.8}$$

$$+ 4.506 \times 10^{-8}\dot{m}^3 \quad \dot{m} \leqslant 200 \text{ Mg/h}$$

From Figure 22.7, it is seen that there is only one valve in the feed section, after the mixing point. Therefore, the only way to reduce the production of phthalic anhydride is to close the valve to the point at which the naphthalene feed is reduced by 50%. Example 22.9 illustrates the consequences of reducing the naphthalene feed rate by 50%.

Example 22.9.

For a reduction in naphthalene feed by 50%, determine the pressures and flows of all streams after the scale-down.

Because it is known that the flowrate of naphthalene has been reduced by 50%, the new outlet pressure from P-201 can be calculated from Equation (22.7). The feed pressure remains at 80 kPa. At a naphthalene flow of 6.41 Mg/h, Equation (22.7) gives a pressure increase of 455.73 kPa, so P_3 = 535.73 kPa. Because the flowrate has decreased by a factor of 2, the pressure drop in the fired heater decreases by a factor of 4 (see Equation [E22.7a]). Therefore, P_5 = 510.73 kPa. Therefore, the pressure of Stream 6 must be 510.73 kPa. The flowrate of air can now be calculated from the compressor curve equation.

There are two unknowns in the compressor curve equation: the compressor outlet pressure and the mass flowrate. Therefore, a second equation is needed. The second equation is obtained from a base-case ratio for the pressure drop across the heat exchanger. The two equations are

$$\frac{P_4}{101.33} = 5.201 + 2.662 \times 10^{-3} \dot{m}_{2,new}^2 - 1.358 \times 10^{-4} \dot{m}_{2,new}^2 + 4.506 \times 10^{-8} \dot{m}_{2,new}^3 \qquad \text{(E22.8a)}$$

$$P_4 - 510.73 = 25 \left(\frac{\dot{m}_{2,new}}{151.47} \right)^2 \qquad \text{(E22.8b)}$$

The solution is

$$P_4 = 512.84 \text{ kPa}$$

$$\dot{m}_2 = 43.80 \text{ Mg/h}$$

The stream table for the scaled-down case is given in Table 22.3. Although it is not precisely true, for lack of additional information, it has been assumed that the pressure of Stream 8 remains constant.

Table 22.3. Partial Stream Table for Scaled-Down Feed Section in Figure 22.7

	Stream							
	1	2	3	4	5	6	7	8
P (kPa)	80.00	101.33	535.73	512.84	510.73	510.73	510.73	200.00
Phase	L	V	L	V	V	V	V	V
Naphthalene (Mg/h)	6.41	—	6.41	—	6.41	—	6.41	6.41
Air (Mg/h)	—	43.80	—	43.80	—	43.80	43.80	43.80

It is observed that the flowrate of air is reduced by far more than 50% in the scaled-down case. This is because of the combination of the compressor curve and the new pressure of Streams 5 and 6 after the naphthalene flowrate is scaled down by 50%. The total flowrate of Stream 8 is now 50.21 Mg/h, which is 30.6% of the original flowrate to the reactor. Given that the reactor was operating at five times minimum fluidization, the reactor is now in danger of not being fluidized adequately. Because the phthalic anhydride reaction is very exothermic, a loss of fluidization could result in poor heat transfer,

which might result in a runaway reaction. The conclusion is that it is not recommended to operate at these scaled-down conditions.

The question is how the air flowrate can be scaled down by 50% to maintain the same ratio of naphthalene to air as in the original case. The answer is in valve placement. Because of the requirement that the pressures at the mixing point be equal, with only one valve after the mixing point, there is only one possible flowrate of air corresponding to a 50% reduction in naphthalene flowrate. Effectively, there is no control of the air flowrate. A chemical process would not be designed as in Figure 22.7. The most common design is illustrated in Figure 22.8. With valves in both feed streams, the flowrates of each stream can be controlled independently. Design of control systems was discussed in more detail in Chapter 18.
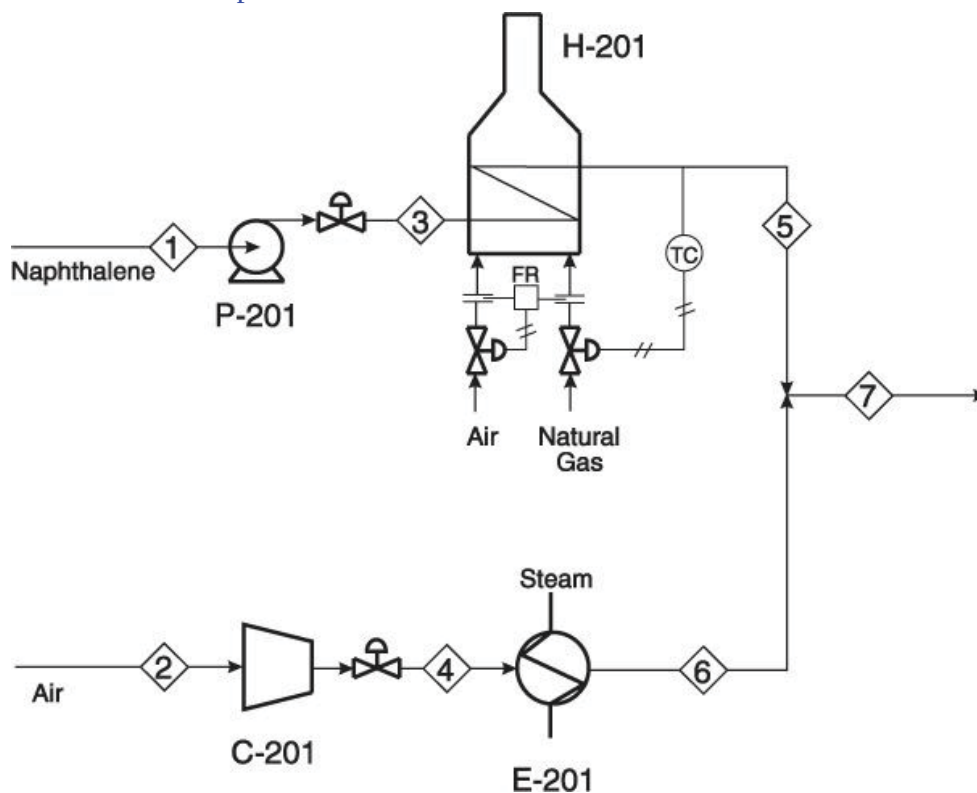
Figure 22.8. Feed Section to Phthalic Anhydride Process with Better Valve Placement than Shown in Figure 22.7

22.5. Summary

In this chapter, it has been demonstrated that performance of existing equipment is affected by other equipment. The input/output models discussed in Chapter 19 suggested this outcome. If the input to a unit is altered, its output is altered. Because the output from one unit is the input to the next, the interaction observed in the case studies in this chapter are expected. It is always important to remember the interaction between equipment performance in a chemical process.

In terms of obtaining numerical solutions for performance problems involving multiple units, examples such as the reactor heat exchanger and the distillation column have been used, in which the adjacent units were analyzed sequentially. Other examples have been used, such as the heat-exchange loop and the feed section in which simultaneous solution of the relationships for multiple units was required. The exact set of calculations necessary and the difficulty of these calculations are specific to each problem encountered. However, the tools developed in Chapter 20, such as determining trends,

base-case ratios, and *T-Q* diagrams, are essential to obtaining desired solutions.

What You Should Have Learned

• The analysis of the previous chapter can be applied to multiple unit operations.

• The general procedure is the same as for single unit operations.

• The input to any unit operation that is the output of another unit operation cannot be assumed; it must be calculated.

References

1. Perry, R. H., and D. W. Green, eds., *Perry's Chemical Engineers' Handbook*, 7th ed. (New York: McGraw-Hill, 1997), 2-61–2-75.

2. Box, G. E. P., *Evolutionary Operation: A Statistical Method for Process Improvement* (New York: John Wiley and Sons, 1969).

3. Kelley, P. E., "EVOP Technique Improves Operation of Amoco's Gas Producing Plants," *Oil and Gas Journal* 71, no. 44 (1973): 94.

Short Answer Questions

1. In the process to produce acetone from isopropyl alcohol (see process description on the CD), heat is supplied to the endothermic reaction using a Dowtherm A (or molten salt) loop and a fired heater. During the start-up of a new plant, it is found that the activity of the catalyst (k_o in the kinetics expression) is less than designed. How would you adjust the flow of heat transfer fluid (up or down) to compensate for this error? Explain your reasoning.

2. In a distillation column, if the bottom pressure decreases, what is the effect on the bottom temperature, the top pressure, the top temperature, and the flooding situation? Assume that all flows, temperatures, and pressures for the utility streams remain constant.

Problems

3. It is possible to generate performance curves for the reactor in Section 22.1. One type of performance curve would have the rate of cumene production (kmol/h) on the *y*-axis, and the percent propane impurity in the propylene on the *x*-axis. There would be lines for different operating temperatures all at constant pressure.

   Assume that the feed to the reactor is 108 kmol/h propylene, 8 kmol/h propane, and 203 kmol/h benzene. The pressure is kept constant at 3000 kPa. Prepare performance curves on the same graph for temperatures from 350°C to 400°C in 10°C intervals. Superimpose on this graph the maximum allowable conditions, which correspond to a steam-side pressure of 4800 kPa. A process simulator should be used to generate points on the performance curves. Other data: reactor volume = 7.89 m³, heat transfer area A = 436 m², overall heat transfer coefficient = 65 W/m²°C.

4. Repeat Problem 22.3 for varying reaction pressure at a constant temperature of 350°C. The pressure range is 3000 kPa to 3300 kPa, in intervals of 50 kPa.

5. For the distillation performance problem in Section 22.2, assume that scale-down occurs while maintaining constant boil-up rate. Determine the conditions in the reboiler, column, and condenser for these operating conditions. A process simulator should be used for the distillation

column.

6. The benzene-toluene distillation column in Section 22.2 must temporarily handle a 25% increase in throughput while maintaining the same outlet concentrations. Determine the operating conditions required if the reflux ratio remains constant. What other factors must be considered in order to determine whether the column can handle this amount of increased throughput?

7. Suggest alternative changes in process conditions or addition of new equipment in the distillation column in Section 22.2 that would allow 50% scale-down. Suggest as many alternatives as you can think of, and discuss the advantages and disadvantages of each.

8. Consider the Dowtherm A loop described in Section 22.3. For the options involving multiple pumps either in series or parallel, determine the configuration that provides the maximum scale-up capacity. What are the temperatures of Dowtherm A, T_3, and T_4, and the maximum percent scale-up?

9. Repeat the analysis in Section 22.3 for the case when cooling water is used in the heat exchanger in place of boiler feed water. The cooling water enters at 30°C and must exit at 45°C.

10. Repeat the analysis in Problem 22.8 using cooling water as described in Problem 22.9.

11. Suggest alternative solutions, not involving increasing the Dowtherm A flowrate, to obtain the maximum scale-up possible in the Dowtherm A loop in Section 22.3. Suggest as many alternatives as you can think of, and discuss the advantages and disadvantages of each.

12. Consider the molten salt loop for removal of the heat of reaction in the production of phthalic anhydride described in Appendix C (on the CD) and illustrated in Figure C.5. It is necessary to scale down phthalic anhydride production by 50%. Estimate the flow of molten salt for the scaled-down case and the temperatures of molten salt entering and exiting the reactor. You may assume that sensible heat effects (energy necessary to heat reactants to reaction temperature) are negligible.

13. Consider the situation illustrated in Figure P22.13. Due to downstream considerations, P_8 is always maintained at 200 kPa. It is known that $P_1 = 100$ kPa and $P_4 = 100$ kPa, and because the feeds come from storage tanks maintained at constant pressure, they are always constant. It is also known that $P_3 = 225$ kPa, $P_5 = 375$ kPa, $P_6 = 250$ kPa, $P_7 = 225$ kPa, and $\dot{m}_1 = 15{,}500$ kg/h. The pump curves are given by the following equations, with ΔP in kPa and $\dot{m}$ in Mg/h:
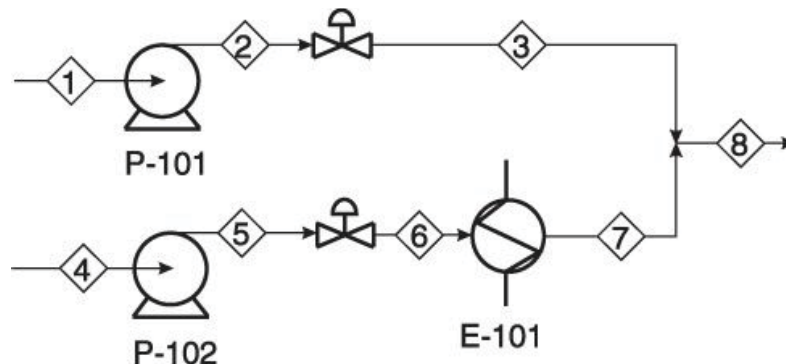

Figure P22.13. Process for Problem 22.13

Pump 1: $\Delta P = 340 - 0.913\,\dot{m} + 0.0535\dot{m}^2 - 0.0101\dot{m}^3$

Pump 2: $\Delta P = 312 - 1.924\,\dot{m} + 0.0302\,\dot{m}^2 - 0.01124\,\dot{m}^3$

a. Calculate $\dot{m}_4$ and ΔP_{23} for this situation.

b. Sketch the pump and system curves for this situation as illustrated in Section 21.2.1. Identify the pressure drop across the valves on the sketch.

14. Refer to Figure P22.14 and the data provided with the figure. The compressor (C-101) exhausts to 101 kPa. Tank TK-102 is controlled to be always 101 kPa. Assume all flows are turbulent.

    a. If the valve is closed, the flowrate from the compressor exhaust (Stream 7) is 100 Mg/h. What is the pressure in the storage tank (TK-101)?

    b. If the pressure of TK-101 is kept constant at the pressure calculated in Part (a), what is the maximum flowrate of liquid (Stream 2) when the valve is wide open (i.e., no pressure drop across the valve)?

Stream or Unit	Pressure Drop
Stream 1	negligible
Stream 2	40 kPa, at 4 Mg/h
Stream 3	30 kPa, flowrate independent
Stream 4	negligible
Stream 5	20 kPa, at 100 Mg/h
Stream 6	negligible
Stream 7	negligible
E-101	70 kPa, at 4 Mg/h
R-101	40 kPa, flowrate independent
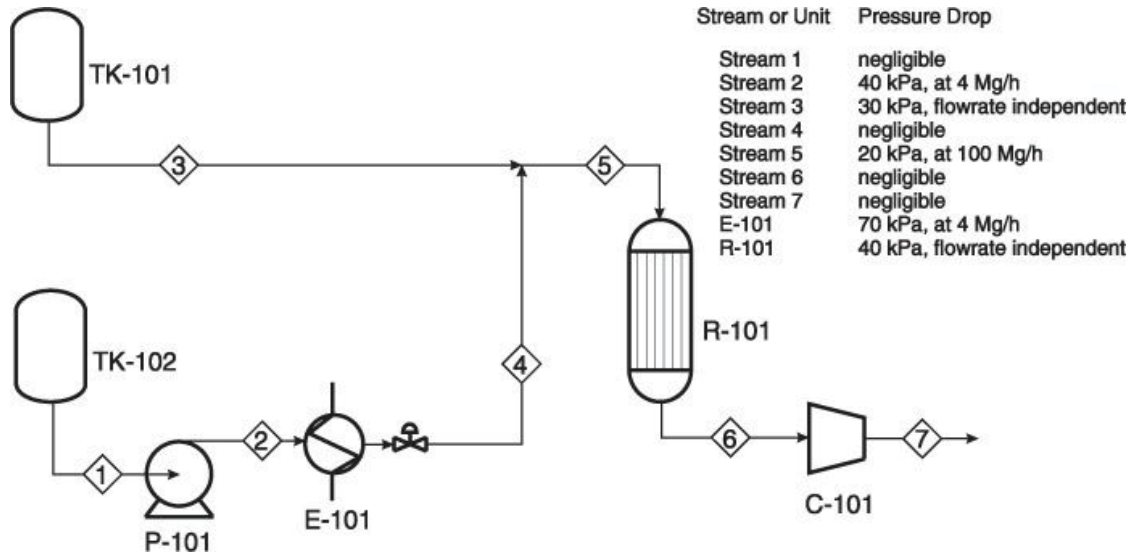
Figure P22.14. Process for Problem 22.14

    c. What is the flowrate of the compressor exhaust (Stream 7) when the valve is wide open?

C-101 compressor curve:

$$\frac{P_{out}}{P_{in}} = 4.015 + 5.264 \times 10^{-3}\dot{m} - 1.838 \times 10^{-4}\dot{m}^2$$

P-101 pump curve:

$$\Delta P = 500 + 4.662\dot{m} - 1.805\dot{m}^2$$

where P is in kPa and $\dot{m}$ is in Mg/h.

15. Consider a process in which a fluid is pumped from a storage tank through a heat exchanger, where it is heated and then pumped into another storage tank. The process flow diagram for this process is shown in Figure P22.15(a), and the pump and system curves are shown in Figure P22.15(b).
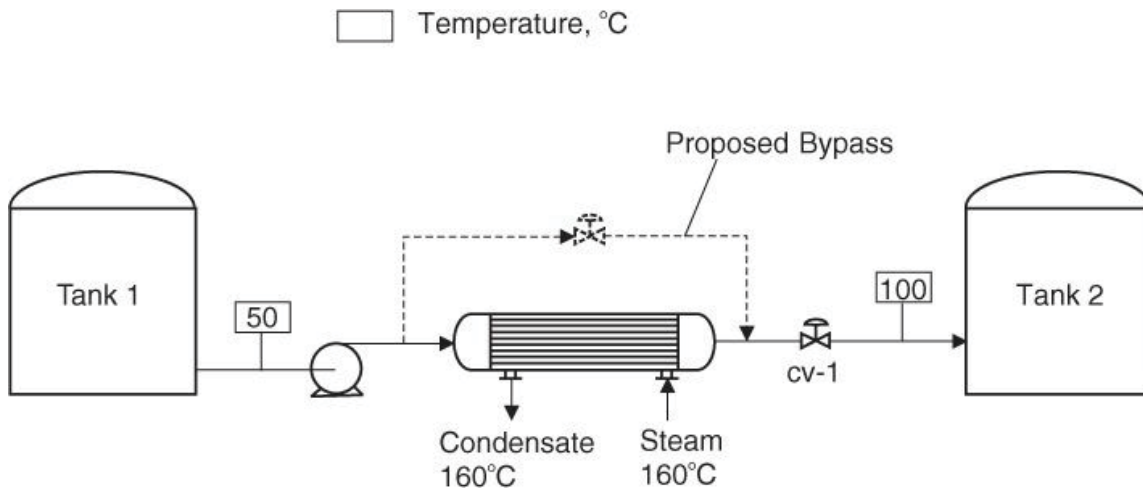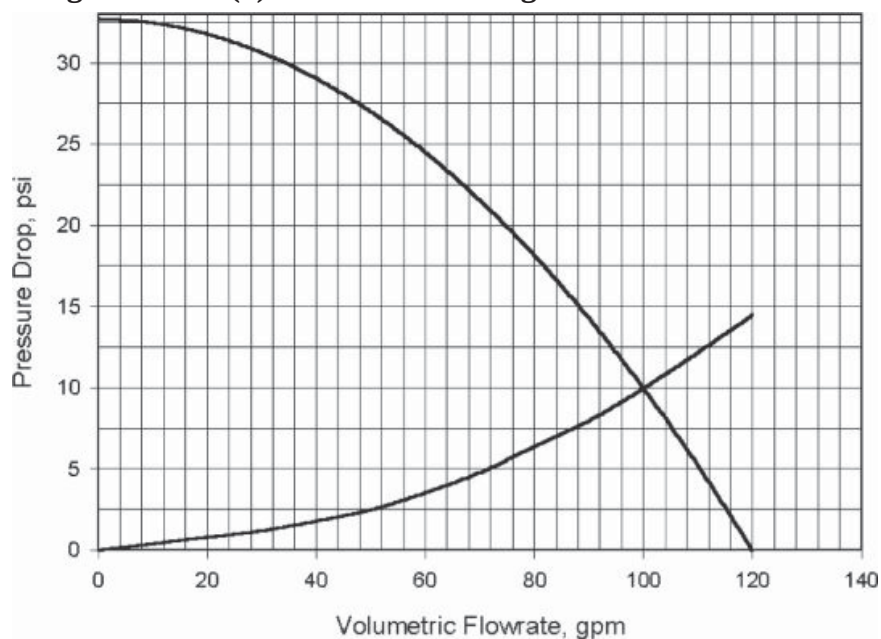
Figure P22.15(a). Process Flow Diagram for Problem 22.15

Pump Curve Equation

$$\Delta P(\text{psi}) = 32.72[1-(\text{vol. flow (gpm)}/120)^2]$$

Figure P22.15(b). Pump Curve and System Curve for Problem 22.15

Currently, the system is operating with the control valve (cv-1) wide open, and the flow through the pump is measured at 100 gallons per minute (gpm). The elevations and pressures of both tanks are equal, and the pressure drop across the heat exchanger is 5 psi, with the remaining 5 psi pressure drop occurring in the piping. It has been noted that the current velocity through the heat exchanger is very high and must be reduced by half. It is proposed to add a bypass line with a valve around the heat exchanger. The valve will be adjusted so that the flow through the heat exchanger is 50% of the current flow. The control valve, cv-1, will remain wide open. For the proposed change, find the following:

a. The pressure drop across the heat exchanger.

b. The total flowrate of fluid passing through the pump.

c. Assume that all the resistance to heat transfer is on the process fluid side and that the temperature of the condensing steam remains constant. Determine the temprature of the fluid entering Tank 2.

16. During the retrofit of a chemical plant, it was decided to place valves and bypass lines around two exchangers placed in series. The flows through each of the bypass lines are controlled by the valves so that 40% of the flow bypasses each exchanger. Because of safety concerns, the bypass lines must be removed. These situations are illustrated in Figure P22.16(a), and the pump curve is in Figure P22.16(b). From the information given in the following table, find the following:

a. For the base case, what is the flowrate through the pump?

b. For the new case, what the maximum flow that can be sent through the two exchangers?

Pressure Profile for Base Case

Location	Pressure (kPa)
1	300
2	290
3	250
4	240
5	200
6	170
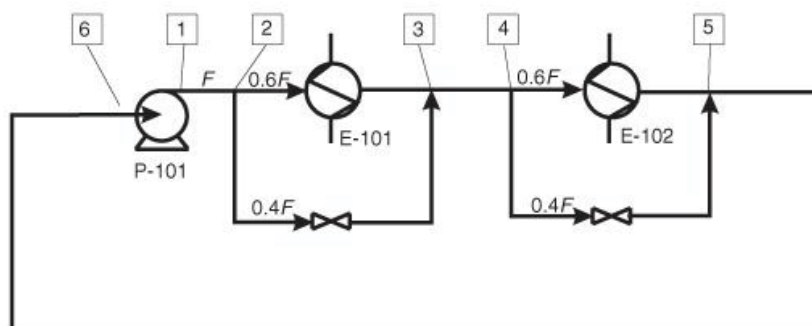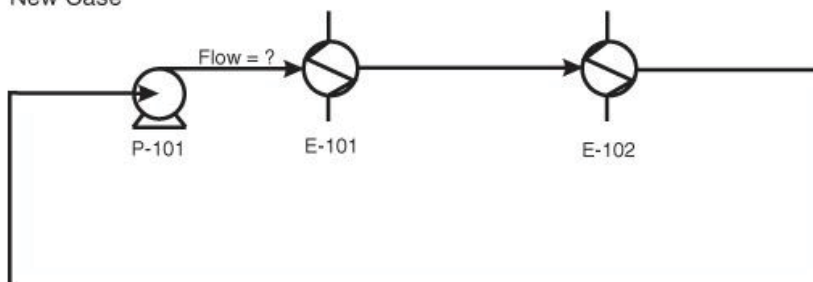
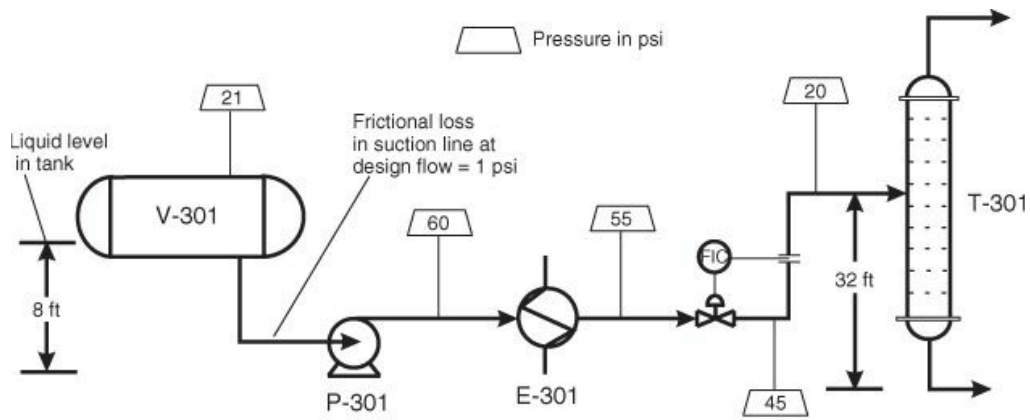**Figure P22.16(a). Process Flow Diagram for Problem 22.16**
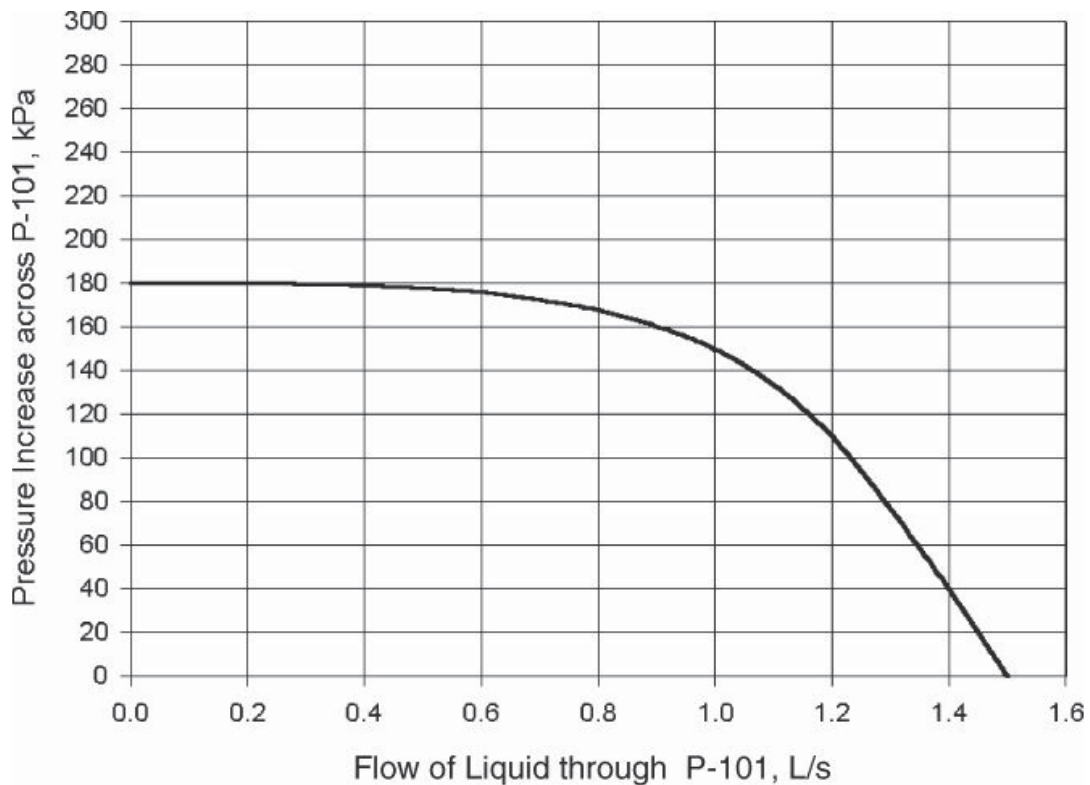
Figure P22.16(b). Pump Curve for P-101

17. The feed system for a distillation process is shown in Figure P22.17(a). The feed to be distilled is an organic liquid stored in V-301 under a nitrogen blanket at a constant pressure of 21 psi. The liquid is pumped through a pump, heat exchanger, and control valve to the distillation column. The pressure profile for the design flow of 50 gpm is shown in the diagram. Also shown are the elevations of the liquid level in the tank and the feed tray in the column. Using the pump characteristics shown in Figure P22.17(b), answer the following questions:

a. What is the maximum flowrate that can be pumped through this system assuming that the feed and destination pressures remain constant?

b. What would the vapor pressure of the liquid have to be at the entrance to the pump so that the pump would just cavitate at the design flow?

c. In order to design the heat exchanger, a maximum design pressure needs to be specified. It has been determined that the maximum pressure to which E-301 can be exposed occurs when the control valve fails closed while the pump continues to operate. What is this maximum pressure?

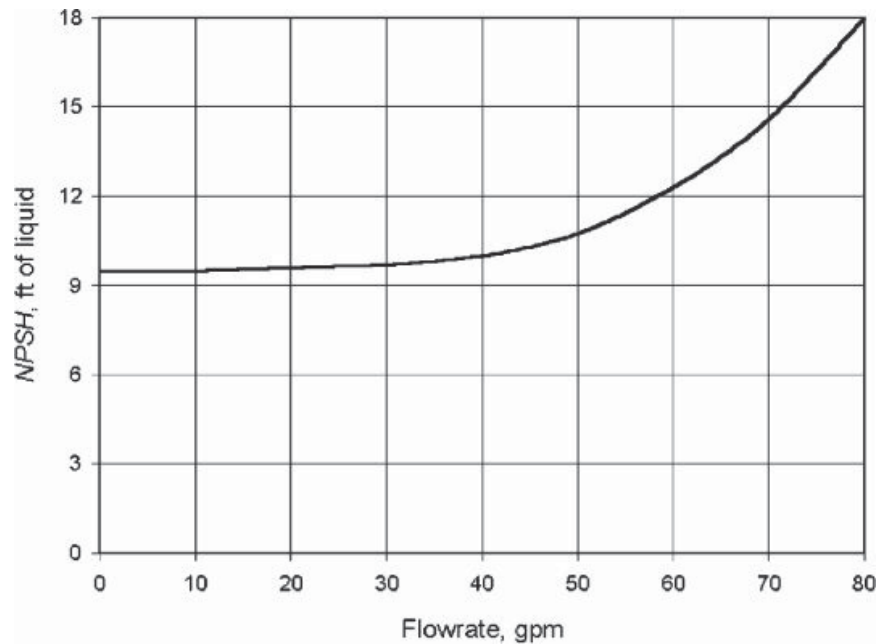Figure P22.17(a). Process Flow Diagram for Problem 22.17
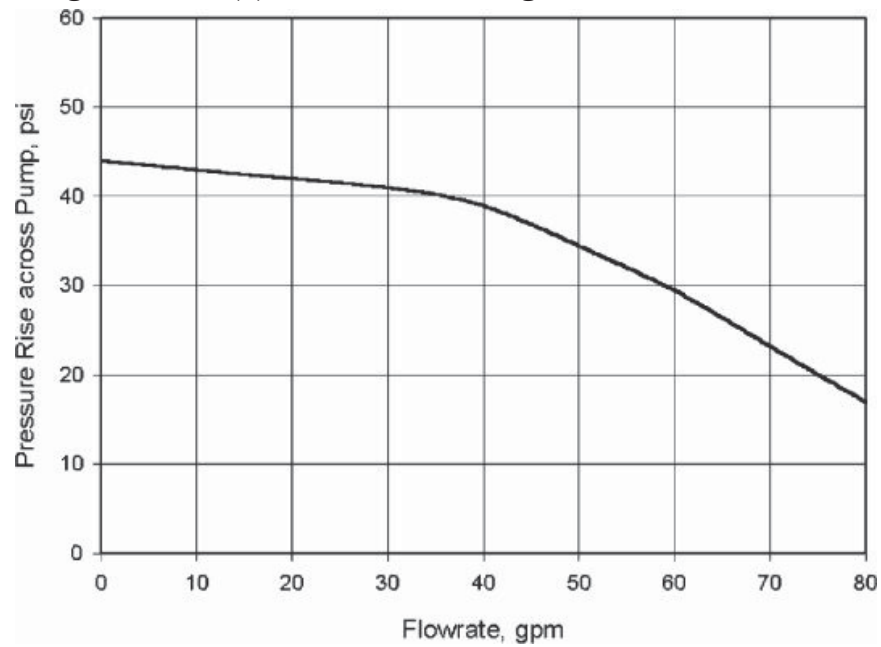
Figure P22.17(b). Pump Characteristics for P-301

18. In the production of acrylic acid, the final step is distillation of an acrylic acid/acetic acid mixture. The acrylic acid, which is the bottoms product, may not exceed 90°C due to spontaneous polymerization above that temperature. Both the acrylic acid and the acetic acid may be considered pure at the bottom and the top of the column, respectively.

    In the current design, a vacuum column with low-pressure-drop, structured packing is used. The pressure drop in the column is 9.5 kPa. The reboiler uses low-pressure steam, but the reboiler feed is equipped with a desuperheater, so the steam temperature is 120°C. There is a control system designed to maintain the temperature difference between the boiling acrylic acid and the steam constant by varying the desuperheater settings. The acrylic acid is reboiled at 89°C. In the condenser, the current design is for cooling water entering at 30°C and leaving at 40°C.

    a. What is the pressure at the bottom of the column?

    b. What is the temperature of the condensing acetic acid at the top of the column?

    c. Based on customer complaints, it is believed that the product is off spec. Investigation shows that the cooling water is flowing at the design rate and that the heat load on the condenser is at design specifications. However, due to a heat wave, the cooling water appears to be entering at 35°C. Could this explain the off-spec product? Support your argument with calculations. What else in the column (temperatures, pressures, etc.) must have changed? What are these new values? Comment on the reboiler control strategy.

    Vapor pressure data:

    For acrylic acid: $\ln P^* (kPa) = 17.31 - \dfrac{5264.4}{T(K)}$

    For acetic acid: $\ln P^* (kPa) = 17.54 - \dfrac{4981.5}{T(K)}$

19. The distillate stream from a distillation column (flowrate 35 m³/h, density of water) is recycled. The destination of the recycle stream is at a location 3 m below the fluid level in the reflux drum. First, the recycle passes through a pump to supply sufficient head to overcome frictional losses

(20 kPa) and reach the pressure of the stream with which it is mixed (250 kPa above distillate stream). Then it is heated from 90°C to 120°C using low-pressure steam at 140°C. Assume that all resistance is on the process-stream side of the exchanger (tube side).

It is desired to scale up the entire plant capacity by 15%. It has been determined that the distillation column can be modified to maintain the recycle stream temperature at 90°C. It has also been determined that, with the required changes in the feed section of the process, the minimum temperature of the recycle stream must be between 119°C and 121°C to satisfy liquid-phase reactor inlet conditions after mixing with fresh feed.

a. Can the temperature criterion be satisfied?

b. Can the flows in this portion of the process be scaled up by 15%? The pump curve is in Figure P22.19.

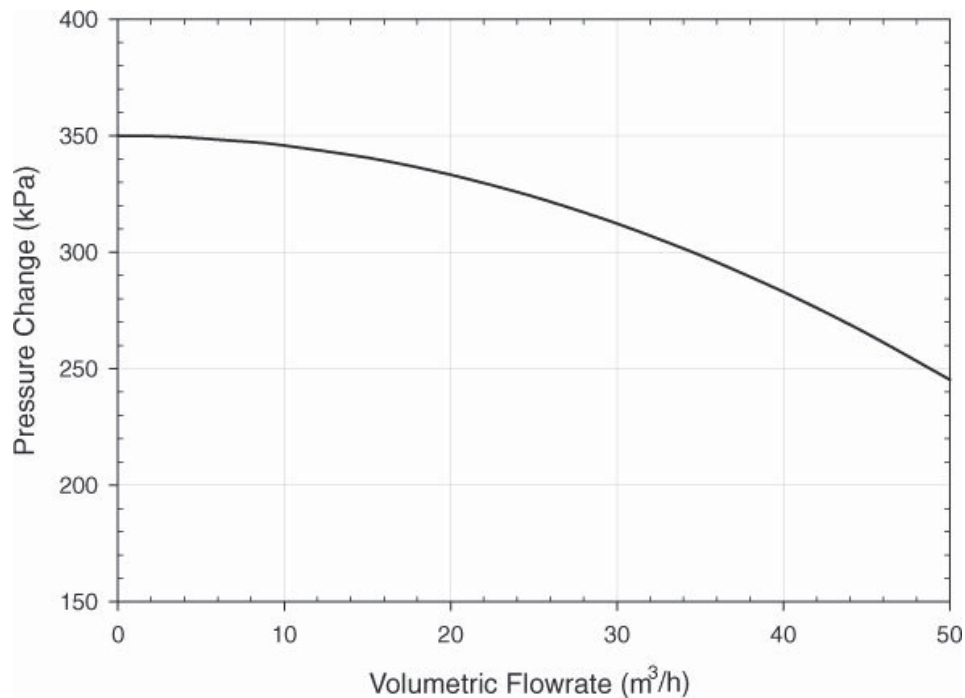Figure P22.19. Pump Curve for **Problem 22.19**

c. With a scale-up of 15% desired, what feature must be included in this distillation column so that column operation can be modified to maintain the recycle stream at 90°C?

d. It is suggested that because a lower temperature of the recycle stream would reduce the reaction rate, if needed, the pumps in the feed section could be used to increase the reactor pressure and hence the reaction rate. What is your opinion of this suggestion?

20. A process heat exchanger has a bypass around it that is normally closed, thus forcing all the process flow through the exchanger. The process fluid (vapor) flows through the shell side of an exchanger, and cooling water flows through the tubes. You may assume that all the heat transfer resistance is on the shell side. During a planned change in operations, it has been determined that additional gas flow will be generated, resulting in an increase of 70% above the current flow. However, the pressure drop across the exchanger would be excessive if all of this flow passed through the exchanger. Therefore, the bypass will be opened so that the pressure drop across the process exchanger increases by only 69%. For this situation answer the following questions:

a. By how much does the flow through the heat exchanger increase?

b. What fraction of the new flow will flow through the bypass?

c. What will be the new overall heat transfer coefficient?

21. You are trying to evaluate whether an existing, idle distillation column can be used for a separation for which it was not originally designed. Answer the following questions about this column:

a. The reflux pump (there are two in parallel) must pump fluid from the reflux drum to the top of the column. The fluid density is 775 kg/m^3. The reflux drum is 6 m above the pump inlet. The top of the tower is 42 m above the pump discharge. Frictional pressure drop for the entire length of pipe is calculated to be 130 kPa. The desired flowrate is 7 L/s. The pump curve and $NPSH_R$ curves are in Figure P22.21. Determine whether this pump can be used for this job and, if so, what configuration is necessary. The frictional pressure drop seems high. Suggest a possible explanation.
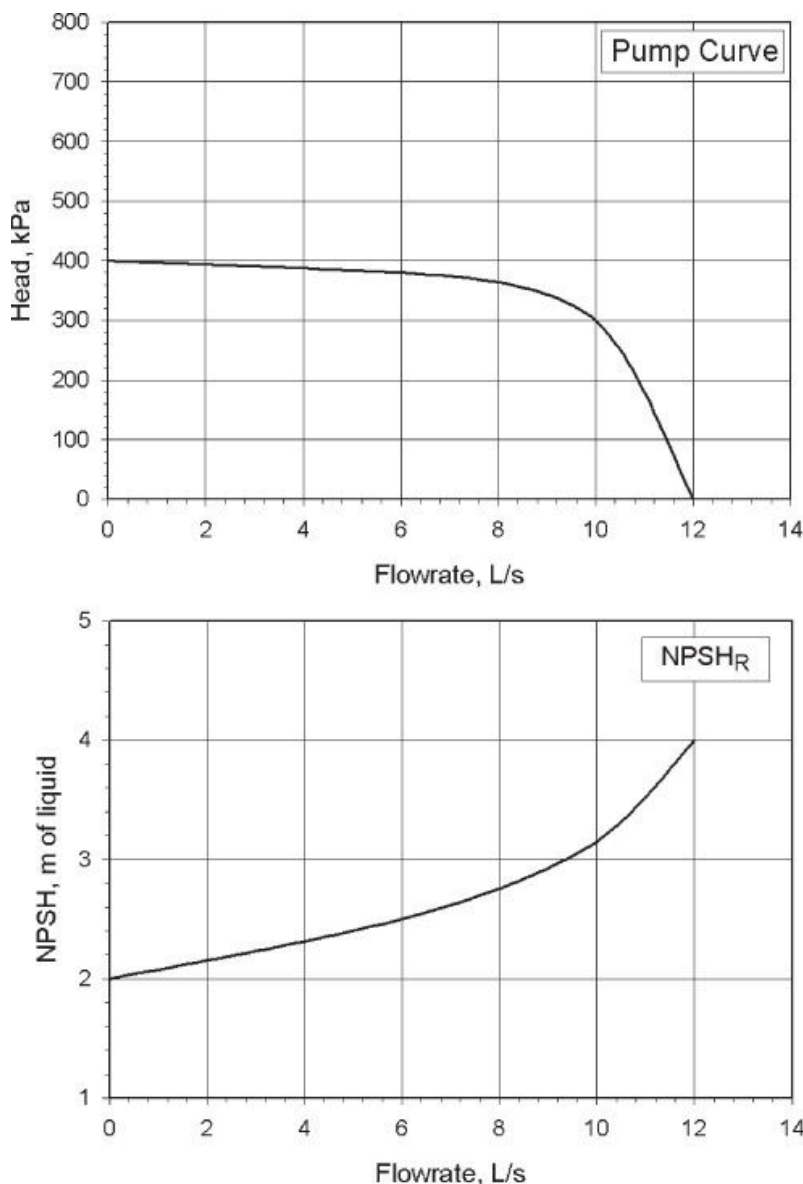

Figure P22.21. Pump and NPSH Curves for Problem 22.21

b. The reboiler is connected to the low-pressure steam line (160°C). However, there is also a

desuperheater. The area of the exchanger is 300 m², and the heat transfer coefficients have been estimated to be 5000 W/m²K and 3000 W/m²K for the boiling side and the condensing side, respectively. The boiling point of the process fluid is 87°C, and the required heat load is 6000 kW. What is the saturated steam temperature required at the desuperheater exit?

   c. List at least three items you would investigate regarding the ability of this column to do the desired separation. For each item, discuss what you would look for or calculate to determine the suitability of the column for this duty.

22. During the design of a new separation system, a reboiler using throttled and desuperheated low-pressure steam (5 barg, T_{sat} = 160°C) was specified. The details of this design are given below:

Duty = 5000 kW

Overall heat transfer coefficient = 1500 W/m²°C

Steam-side coefficient = process-side coefficient

$T_{process}$ = 70°C

T_{steam} (after throttling and desuperheating) = 110°C

The heat transfer coefficient for the boiling process fluid as a function of the temperature difference between the wall and the process fluid for this design is given in Figure P22.22.
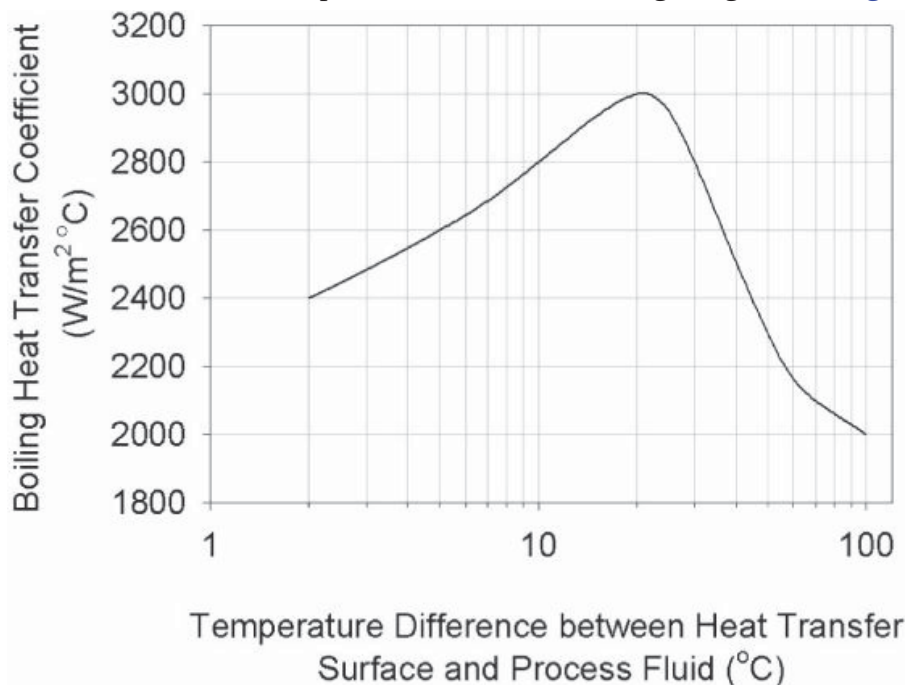

Figure P22.22. Boiling Heat Transfer Coefficient as a Function of Temperature Difference

   a. During the construction of this new separation system, the throttling valve and desuperheating unit have been omitted. By how much will the new reboil rate change (magnitude as a percent of the base case), assuming that the column pressure and temperature do not change?

   Hint: You may assume that the steam-side heat transfer coefficient does not change but the process-side heat transfer coefficient will change. *This means that the wall temperature will also change, and the solution to the problem will require trial and error.*

   b. If the column were allowed to operate without the desuperheater and throttling units and no adjustments or changes were made to the column operation, how would the pressure in the

column change? This answer should be consistent with your answer to Part (a).

23. The bottoms product from a distillation column is pumped through two heat exchangers to recover energy and is then sent to a storage tank, which is at atmospheric pressure. The system is shown in Figure P22.23(a).
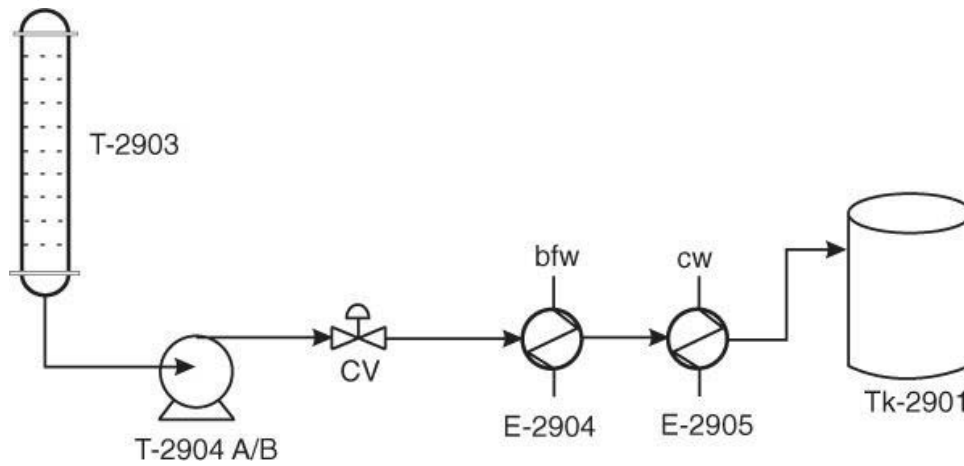

Figure P22.23(a). Process Flow Diagram for Problem 22.23

The pump curve for P-2904A/B is shown in Figure P22.23(b). At design conditions, the flow through the pump is 5 L/s, the pressure in T-2903 is 200 kPa, and the frictional loss (for the process fluid) through the piping and heat exchangers, but excluding the control valve, is 250 kPa. You may assume that the difference in height between the liquid level in T-2903 and Tk-2901 is negligible. For this system, answer the following questions:

a. What is the pressure drop across the control valve at design conditions?

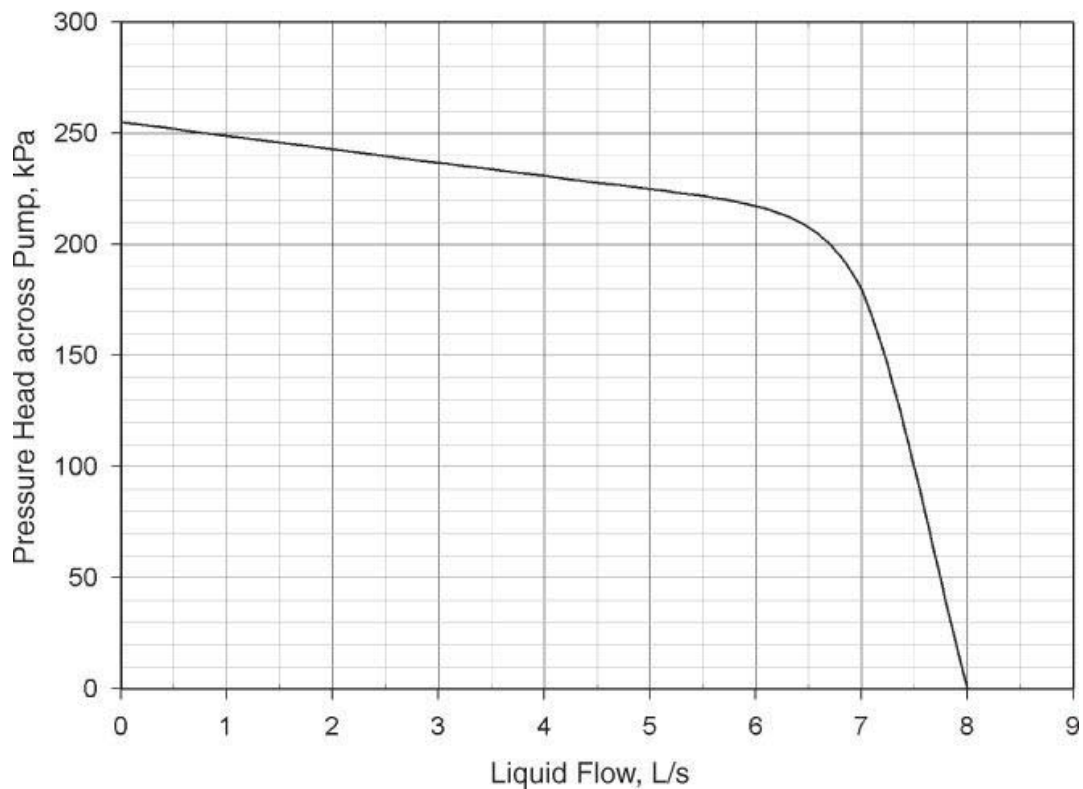b. Under design conditions, the 250 kPa frictional pressure drop is distributed as shown in Table P22.23.

Figure P22.23(b). Pump Curve for Problem 22.23

Table P22.23. Pressure Drops in Pipe Sections

Equipment or Pipe	Pressure Drop at Design Flowrate (kPa)
Suction piping	5
Piping from P-2904 to CV	10
Piping from P-2904 to CV	10
E-2904	50
Piping from E-2904 to E-2905	25
E-2905	50
Piping from E-2905 to Tk-2901	100

  c. What is the pressure of the liquid entering E-2904 and E-2905 at a scaled-up condition, which is 150% of the base-case flowrate?

24. Two heat exchangers of equal area have been designed to heat a certain process fluid from 50°C to 100°C using condensing saturated steam at 150°C. The designs of the two exchangers are different and result in different pressure drops and different overall heat transfer coefficients. Pilot-test results for the two heat exchangers operating individually are given in Table P22.24 for a single set of operating conditions.

Table P22.24. Heat-Exchanger Pilot-Test Results

Process Variable	Exchanger 1	Exchanger 2
Mass flow of process liquid (tube side)	2.0 kg/s	1.5 kg/s
Inlet temperature of liquid	50°C	50°C
Outlet temperature of liquid	100°C	100°C
Process liquid C_p	1000 J/kgK	1000 J/kgK
Steam temperature (shell side)	150°C	150°C
ΔT_{lm}	72.13°C	72.13°C
Area	6.93 m²	6.93 m²
Condensing steam heat transfer coefficient	1000 W/m²K	1000 W/m²K
U	200 W/m²K	150 W/m²K
Tube-side pressure drop	40 kPa	25 kPa

In the plant, these two exchangers are now to be piped in parallel, and at the current operating conditions, the pressure drop across both heat exchangers is measured at 52 kPa.

Assuming that the physical properties and inlet temperature of the process stream (in the tube side) are the same as given in Table P22.24 and that condensing steam is used in the shell side, answer the following questions:

a. What is the mass flowrate of process fluid through Exchanger 1?

b. What is the mass flowrate of process fluid through Exchanger 2?

c. What is the new overall heat transfer coefficient for Exchanger 1?

d. What is the new overall heat transfer coefficient for Exchanger 2?

e. If the process exit stream from Exchanger 1 is to be maintained at 100°C, what must be the temperature of the condensing steam that is fed to this exchanger?

f. If the process exit stream from Exchanger 2 is to be maintained at 100°C, what must be the temperature of the condensing steam that is fed to this exchanger?

g. Under another set of conditions, if the two exchangers were piped in parallel and the flow measured through Exchanger 1 was 3.1 kg/s, what would the flow be through Exchanger 2?

25. Consider the system illustrated in Figure P22.25(a). The process is to be scaled up from a design flowrate of 120 m³/h. The fluid is a liquid, and it is to be fed to a reactor. The heating medium is condensing, medium-pressure steam throttled and desuperheated to saturated steam at 180°C. In the heat exchanger, the resistances are such that the process-side resistance (tubes) is three times that of the condensing steam (shell). The pump and system curves are shown in Figure P22.25(b).

a. What is the maximum percent scale-up possible in this system and the corresponding outlet temperature of E-101?

b. If it is necessary for the minimum reactor feed temperature to be 153°C, what is the maximum possible scale-up?
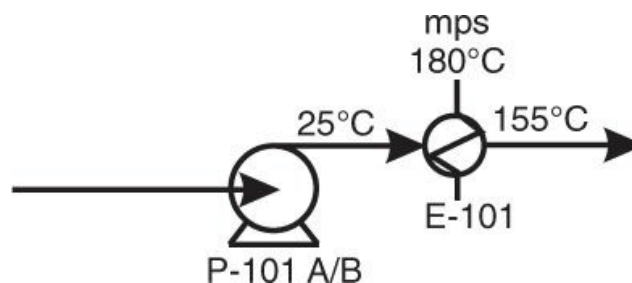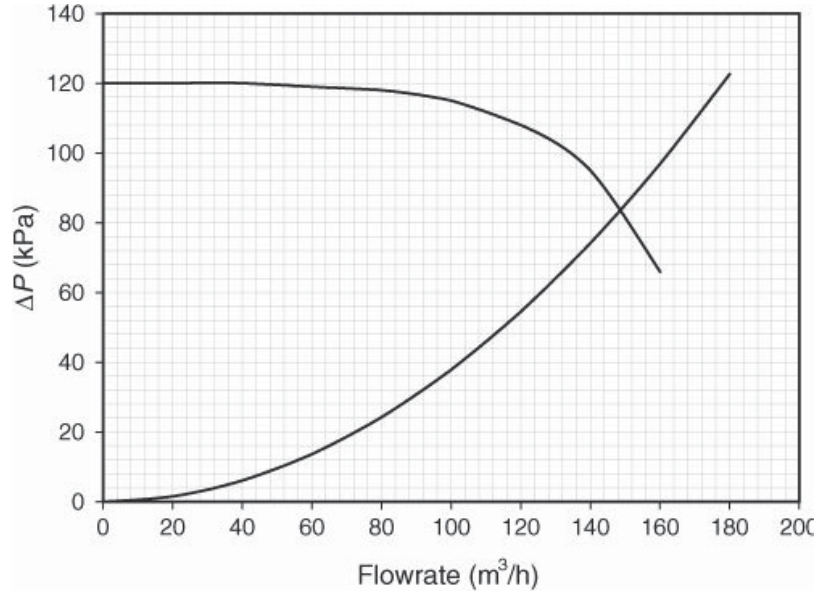
Figure P22.25(a). PFD for Problem 22.25

Figure P22.25(b). Pump and System Curves for P-101 A/B in Problem 22.25