If 50% of the n-hexane leaves in the vapor product, what is the ratio $v_{n\text{-hexane}}/l_{n\text{-hexane}}$? If you know P, can you estimate T and vice versa? You should note that, for each guess of the relative volatility, the flash computation asked for here does not require iteration.)

c. Assume that you wish to design a column to separate the n-heptane from the remaining three species as the first column in the sequence selected to carry out the complete separation. Assume the feed and both products are bubble point liquids. Estimate the minimum reflux ratio for this column. Is the method you used justified for computing this minimum reflux? Explain.

d. Develop the condenser and reboiler heat duties for the column for pressures of 1, 5, 10, and 20 atm. Plot heat duties versus the condenser temperature for this column. Do you notice anything special about this plot?

e. Would you use this method for a column to separate acetone from ethanol (see exercise 1)? Explain.

7. Enumerate all the simple sharp separation sequences possible for separating a mixture of $ABCDE$ into products AC, BE, and D given the following three separation methods:

   • Method m1: Component volatility order $ABCDE$
   • Method m2: Component volatility order $CBADE$
   • Method m3: Component volatility order $BCED$

   For you to use method 3, species A may not be present.

8. Estimate the minimum reflux using Underwood's method for separating the following mixture into the products indicated.

Species	Feed	Recovery in Distillate
n-pentane	20%	100%
n-hexane	50%	99.5%
n-heptane	30%	0.2%

9. Consider the mixture in Table 11.10. Using Underwood's equations, compute the minimum reflux to recover 90, 95, 99, and 99.9% of the key components in their respective products for the following separation problem. Species C and D are the key components.

TABLE 11.10   Mixture for HW Problems

Species	Relative Volatility	Feed Flow, mol/s
A	2.7	10
B	2	5
C	1.5	40
D	1	15

10. Again consider the mixture in Table 11.10. Underwood's equations can be used for computing the minimum reflux when the key components are not adjacent in the

separation. Let the light key be species A and recover 99.5% of it in the distillate. Let the heavy key be species C and recover 99% of it in the bottoms product. Find two roots for the first Underwood equation: the one that lies between A and B and the one that lies between B and C. Write the second of the Underwood equations twice, once using the AB root and once using the BC root. You should have two linear equations in two unknowns: the flow d_B and in the minimum vapor flow, V_{min}. Solve these two equations for these flows.

11. Let the light key remain the same as in the previous problem. Let the heavy key be species D. Recover 98% of it in the bottoms product. What is the minimum vapor flow for this column? Problem 10 describes how to solve this problem.

12. Discover the best sequence among those possible for the following problem based on minimizing the total of the estimated vapor flows in the columns.

Species	Relative Volatility	Amount kmol/hr
A	2	10
B	1.5	20
C	1.2	10
D	1	60

Is the answer consistent with any of the heuristics in Table 11.8? Explain.

Suppose that species B is very corrosive. Estimate the extra cost in terms of added vapor flow for following the "dangerous or corrosive species" heuristic.

13. Consider again the mixture consisting of 35 mole % n-heptane, 30% n-hexane, 10% isobutane, and 25% n-pentane. Using Eq. (11.5), estimate marginal vapor rates and determine which of the possible sequences constructed from simple two product columns are likely to be the best. Would you expect this heuristic to give the right answer here? Explain.

14. Find the best distillation-based separation sequence if the following data hold for marginal vapor flows using a branch and bound search. The components behave relatively ideally.

	A	B	C	D	E
A/B	—	—	100	1	1
B/C	1	—	—	1	1
C/D	1	100	—	—	1
D/E	1	1	100	—	—

Prove that you have the best answer by listing the total marginal vapor flows for all sequences.

15. You wish to separate a mixture of species A, B, and C using distillation. These species have fairly ideal vapor/liquid equilibrium behavior, having relative volatili-

ties of 4.0, 2.0, and 1.0 respectively. The flowrate of species C in the mixture is 1 kmol/hr. Estimate the flowrates of A and B in the feed such that you would be indifferent to choosing between the direct (A/BC, B/C) and the indirect (AB/C, A/B) sequences for separating them.

16. Consider separating the mixture in Table 11.11 into four pure component products.

TABLE 11.11   Feed Flow for Exercise 16

Species	Feed Flow, mol/s
n-pentanol	10
isobutanol	5
n-hexanol	40
n-heptanol	15

   a. Using Underwood's equations, find the sequence having the lowest total for the minimum vapor flows in each of the columns in it.
   b. Use the marginal flow estimator given by Eq. (11.5) and find the sequence having the lowest total for the minimum vapor flows in the columns in it.
   c. Compute the marginal flows using the results from part a and compare them to part b.

17. Using the heuristics in Table 11.8, find a reasonable separation sequence for the feed in Table 11.11. If you have done the previous problem, how does this answer compare?

18. Using the heuristics in Table 11.8, propose separation sequences for the following problem.

   Separate a mixture of six components $ABCDEF$ into products A, BDE, C, and F.

   Use either of two methods in developing your sequences

   • Distillation, method I   Component volatility order $ABCDEF$
   • Extractive distillation, method II   Component volatility order $ACBDEF$

Component amounts

   • A: 4.55 kmols/hr, B: 45.5, C: 155.0, D: 48.2, E: 36.8 and F: 18.2.

Relative volatilities of the key species

   • Method m1: A/B 2.45, B/C 1.55, C/D 1.03, E/F 2.50
   • Method m2: C/B 1.17, C/D 1.70

19. Show that the direct sequence is the correct one for the following problem. Note that all the volatility ratios for adjacent species, $\alpha_{i,i+1} = r$, are equal to 1.2 here.

Species	Relative Volatility	Amount kmol/hr
A	$1.2^3 = 1.728$	1
B	$1.2^2 = 1.44$	1
C	$1.2^1 = 1.2$	1
D	$1.2^0 = 1$	1

20. Show that the result for the previous problem is general for any ratio $\alpha_{i,i+1} = r$ and not just for $r = 1.2$.

21. List the total number of instances of *extra species* present for each of the possible sequences when splitting an 8-component feed mixture into 8 relatively pure component products. Which sequence has the fewest number of extra species overall? Discuss the implications of having the fewest total number of extra species on the marginal vapor flow.

There is an heuristic that says that a column should attempt to split each mixture in a separation process into roughly equal parts. Explain how the above observation on extra species may support this heuristic.

HEAT INTEGRATED DISTILLATION PROCESSES

12

In this chapter we combine the topics of the last two chapters to look at the heat integration of systems of distillation columns. We shall also look at special column configurations that feature intercooling and interheating as well as columns that have side strippers and enrichers.

12.1   HEAT FLOWS IN DISTILLATION

12.1.1   A Base Case (Andrecovich and Westerberg, 1985)

Distillation columns require heating for the reboiler and cooling for the condenser. Unfortunately, but, not surprisingly, the reboiler, always hotter than the condenser, cannot directly use the condenser heat. Columns are *heat integrated* if heat removed from one is used to provide heat for another. Often, we have to adjust the temperature levels for the columns involved so they can be integrated, but, fortunately, we can increase or decrease the operating temperatures for a column by simply increasing or decreasing its operating pressure.

Columns can be viewed as devices that degrade heat to carry out separation. They receive higher temperature heat into their reboilers and expel lower temperature heat from their condensers. Higher temperature heat should, and had better, cost more per unit of heat than lower temperature heat. In an ideal world, we would buy utilities at just the temperature needed, paying a price for them that reflects their temperature. In such a situation, passing heat from one column to another would probably not be economic.

However, most utility systems for processes provide heat at only a few fixed temperature levels—for example, from high, medium, and low pressure steam at 350, 275,

408

and 200°C, respectively. Suppose we have a column that has a condenser temperature of 50°C at one atmosphere (hot enough to pass the heat into cooling water) and a reboiler at 90°C. We would like to use 100°C utility heat, except there is none. We find we must use 200°C steam. It could prove economical to use this same heat to run one or more other columns before it passes through this column. This column will degrade the heat passing through it by only about 40°C (90°C less 50°C) plus the sum of the temperature differences used as driving forces in its reboiler and condenser, say another 20 to 30°C.

It is also possible that we could exchange heat with other streams in the process.

When heat is degraded and passed to another part of the process to degrade it further, there is a cost. The temperature driving forces for heat exchange will become smaller. If small enough, the heat exchangers for a column can cost more to purchase than the column itself. (Nothing is free.) The following ideas illustrate those instances when heat integration might be attractive because of the potential utility savings. Only these ideas need to be investigated as they are the only ones that could produce a savings that can pay for the extra exchanger area required.

Both the first and second laws are at work here. We would like to reduce the use of utilities by reusing heat (first law savings). However, the heat is degraded each time we use it (second law cost). Because of the large temperature drops available when using only a few temperature levels for utilities, we are often forced to pay for the large temperature drops whether we use them or not. Forced to have them, we should try to use them.

In order to explore these possibilities, we need to understand and be able to compute the heat flows in columns. That is the purpose of this section.

We start by considering a *base case* column, one that we shall use to compare the operation of all others. Assumptions for this base case column are:

- Feed and products are all liquids at their respective bubble points (i.e., they are liquids at their boiling point).
- Internal reflux and reboil flow rates are large relative to feed and product flow rates.

A heat balance around the column gives

$$h_F(T_{f,\text{bub}})F + Q_{\text{reb}} = h_D(T_{D,\text{bub}})D + h_B(T_{B,\text{bub}})B + Q_{\text{cond}}$$

With the above assumptions, the terms Q_{reb} and Q_{cond}, which involve latent heats, are very large compared to the remaining terms which involve only differences in sensible heats. Thus, we can write

$$Q_{\text{reb}} \approx Q_{\text{cond}}$$

A column for the base case degrades approximately $Q = Q_{\text{reb}} \approx Q_{\text{cond}}$ units of heat from T_{reb} to T_{cond}. In Figure 12.1 we sketch this base case as horizontal heat source and heat sink lines of width Q on a plot of T versus *Heat*. We can think of the horizontal lines being joined top to bottom to form a box *for this case*. While tempting and something we have done often, we will not show columns as boxes because the duties are often not equal for a column, for example, when the feed is dewpoint vapor.
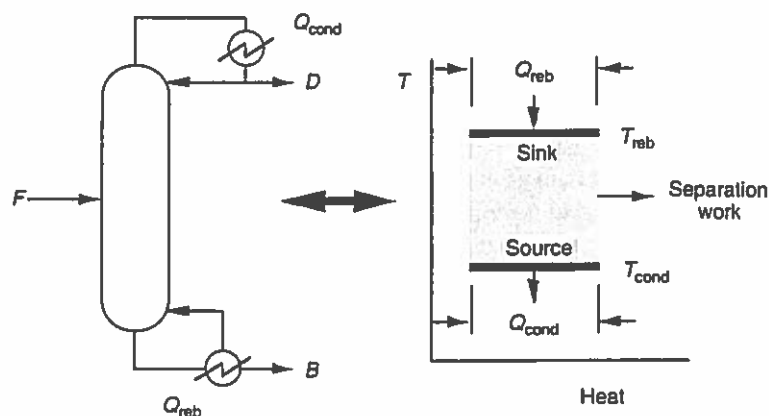
FIGURE 12.1 Base case heat balance for column—the T-Q diagram for distillation.

OBSERVATIONS ON T-Q DIAGRAM

The following observations for a column come from having carried out computations for many different examples. Most experience is with relatively ideally behaving species.

- Higher pressure $\rightarrow$ higher temperature operation $\rightarrow$ *both* more heat required and a larger temperature drop across column—that is, the box gets larger in both dimensions.

Intuition would suggest that more heat should be needed as higher pressures generally lead to smaller relative volatilities between the species; at least that is the experience with normal hydrocarbons. Thus, more reflux would be required. One's intuition probably would not suggest that the temperature drop should also increase, but it does.

- Having other species present typically increases both the heat duties and the temperature drop across the column.

We saw in the previous chapter that there is an added vapor flow when other species are present. The temperature drop increase is also expected as having D present for the B/C split will increase the bubble point for the reboiler (CD rather than for C alone).

COMPUTING REBOILER AND CONDENSER DUTIES

The following is a recipe to estimate condenser and reboiler duties for a column. Because of the effects of composition on enthalpies, it cannot be exact.

- Estimate the minimum reflux/reboil ratio required for column.
- Select a reflux/reboil that is, say, 1.2 times as large as the minimum needed.
- Multiply the heat of vaporization for the distillate/bottoms times the reflux/reboil used.

SYSTEMS OF HEAT INTEGRATED COLUMNS

To indicate the type of thinking involved in heat integrating columns, we consider the following example where we shall use NO numbers. The T versus Q representation for heat flows in columns will allow us to gain insights into the design for this problem none-the-less.

EXAMPLE 12.1

Split the following mixture of components.

Species	Amount	Ease of Separation
A	lots	
		difficult
B	moderate amount	
		very easy
C	moderate amount	
		very very difficult
D	lots	

Figure 12.2 sketches the T-Q flows for each of the separations for this example. Separating C from D is difficult, indicating they have close boiling points. The temperature drop across the column is, therefore, small, but the amount of heat required is very large, as shown. On the other hand, separating B from C is easy. Here the normal boiling points will be very different—that is, there is a large temperature drop, but the heat needed is very little. Finally, splitting A from B is somewhere in between.

We make the following observations based on our understanding of how distillation processes work.

- C/D should be done without other species present—other species will enlarge the amount of heat required for a column that has a large heat requirement already.
- B/C should be done without other species present. This preference conflicts with the previous one. With a large temperature drop, it is difficult to heat integrate this column with others and still be within the allowed utility temperatures. The potential benefits of reusing heat passing through this column are greatly reduced.
- The C/D split could conceivably be done in two columns that are heat integrated to reduce the utility consumption (carrying out the same separation in two columns and heat integrating them is termed multi-effect distillation—for reasons that hopefully are obvious.
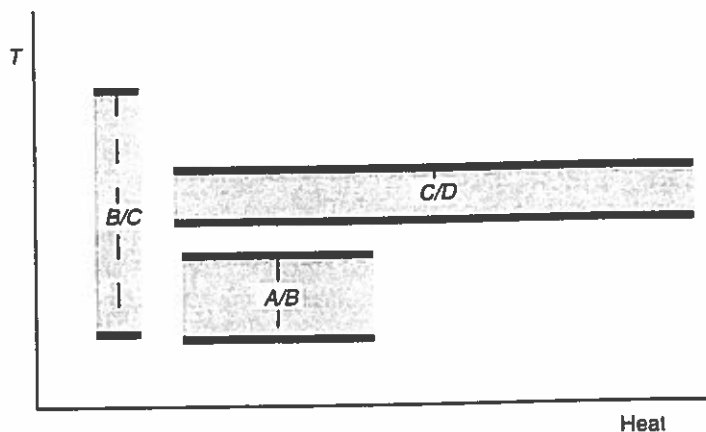

FIGURE 12.2 T-Q heat flows for example splits.

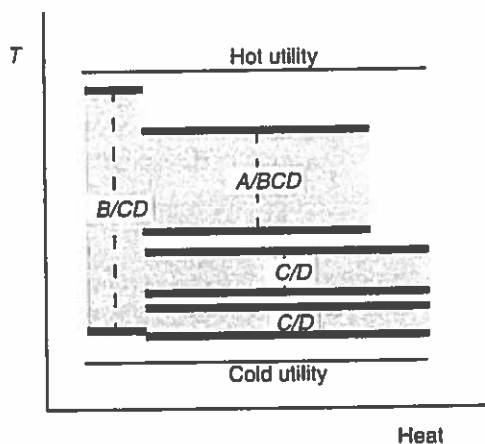We select the candidate design in Figure 12.3 based on these assumptions for the process.

FIGURE 12.3   Heat integrated design for separation problem. A box placed vertically above another implies heat passes from the condenser of the column corresponding to the upper box into the reboiler of the column for the lower box.

The following give the reasons for this design.

- C/D is done without other components present.
- The box for C/D was split in a manner that both parts will have same width in the final design. Thus, the heat required by one is exactly the heat given up by the other.
- The two boxes for the C/D split are operated at the coldest temperature possible to reduce the dimensions for them. Their width impacts directly the amount of the utilities that are consumed.
- The split B/C is done with fewest other components possible; if others have to be present, we choose to have heavy species as they have a smaller effect on added heat duties; here we must have D present if the C/D is split is done with no other species present.

Note that the dimensions for the heat flows and temperature drops reflect that the columns are operating at different conditions than in the previous figure (different temperature levels, different components present).

12.1.2   Intercooling/Heating

An interheated and/or intercooled column is one in which heat is added and/or removed from trays within the column (the following analysis is from Terranova and Westerberg, 1989). In our previous columns, all heat was added to the reboiler and removed from the condenser. Questions we might ask are:

- Why use intercooling or interheating?
- Is more or less heat required?
- What are the costs?

We start by examining a binary separation for which we can construct a McCabe-Thiele diagram. The column in Figure 12.4 has two envelopes for which we might write component material balances at the top of the column, one above the intercooler and one below.

The operating lines for each are a result of writing component material balances:

$$y = \frac{L^I}{V^I} x + \frac{L^I}{D} x_D$$

$$y = \frac{L^{II}}{V^{II}} x + \frac{L^{II}}{D} x_D$$
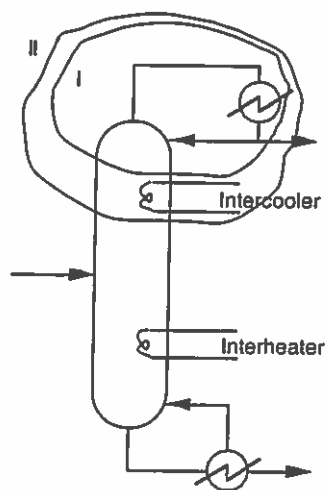

FIGURE 12.4   Material balance line for intercooling.

Since the top product is the same for both envelopes, both operating lines must go through the same point $[x_D, x_D]$ on the 45-degree line. The only thing that can vary is the slope for each of them, which can be written in the following form for both.

$$\text{slope} = \frac{L}{L+D} = \frac{1}{1 + \dfrac{D}{L}}$$

Intercooling will cause L to be larger for envelope II, and therefore its slope, by the above, will be larger (i.e., larger L implies a smaller denominator implies a larger quotient). As a point of interest, we also note that since $V = L + D$ for both cases, V must also be larger for envelope II.

Figure 12.5 illustrates the McCabe-Thiele plot for a binary separation with intercooling and interheating. In the top part of the column, not removing enough heat from the condenser to run the column leads to an operating line with too small a slope to reach the bottom operating line before it crosses the equilibrium curve. Removing heat partway down pivots the operating line downward to give it a steep enough slope. We see similar behavior for the bottom of the column, where not placing enough heat into the reboiler leads to an operating line that is too steep to reach the upper line before it crosses the equilibrium curve. Shown also are the stages required for this column. We note that the temperature for a column increases as we march down it, so $T_1 < T_2 < \cdots T_9$.

We step along the first operating line (stages 1 and 2), getting warmer as we go, until we step over the intercooler, where we move to the lower operating line. With an intercooler we can remove less heat in the condenser than needed for an ordinary column because we can "rescue" the operating line and move it down by intercooling before it
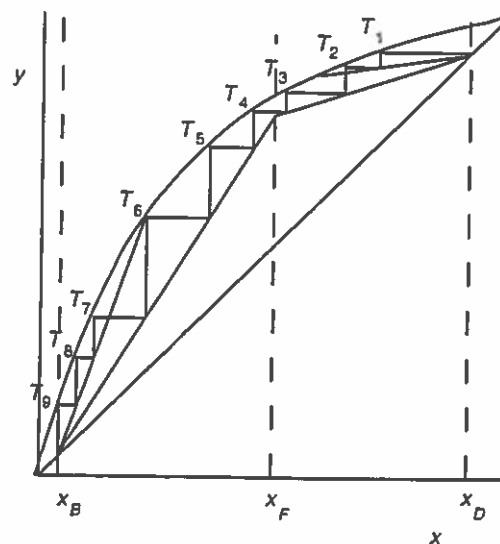

FIGURE 12.5   McCabe-Thiele plot for a column with intercooling and interheating.

pinches with the equilibrium line. The intercooler removes heat partway down the column and, therefore, at a higher temperature than the condenser.

We can also observe that the minimum reflux requirement for the column dictates the slope for the *second* operating line only, irrespective of whether we have an intercooler. We can therefore argue that the total heat removed, which dictates the slope for the second operating line, has not changed. We have only altered the conditions at which some of the heat has been removed.

Answers to questions about intercooling that we asked earlier are now more evident.

1. Intercooling allows us to remove only part of the heat in the condenser. At a warmer temperature (between T_2 and T_3 in our example), we then remove the remaining heat. By a similar set of arguments, interheating allows us to inject only a part of the heat into the reboiler where the column is hottest. At a lower temperature we then inject the remaining heat needed to run the column.
2. If we do not move the operating line for envelope II and insist on producing the same products, then the same total amount of heat is removed and injected as for a normal column, and we find that we require more trays (as the steps along the operating line for envelope I are smaller). We also need to purchase the heat exchanger equipment, and, if we use the same utilities, it will have a smaller temperature driving force and thus require more heat transfer area. The heat exchanger equipment will almost certainly be more expensive.
3. If we have a column with a fixed number of trays (as we would for the retrofit case) and we leave the operating lines to have the same slope for envelope II, then the column will give a poorer separation. To accomplish the same separation, we have to increase the reflux we use in the column, moving the operating line for envelope II, and possibly for envelope I, closer to the 45-degree line. We would almost certainly need more heat exchanger equipment.

For 2 above, one gains on the second law—i.e., one can remove heat at hotter temperatures and inject it at lower temperatures, stays even on the first law—i.e., the column uses the same amount of heat, and finally one has to spend more on equipment as more trays and exchanger equipment are needed. For 3, one again gains on the second law but either loses on the separation accomplished or loses on the first law.

HEAT FLOWS FOR INTERCOOLED/INTERHEATED COLUMNS

The T versus *heat* diagram should have the shape shown in Figure 12.6, where the darkened lines are the heat in and out lines. The outer box is the T versus *heat* diagram for a column without interheating and intercooling. The impact of interheating and intercooling is to notch the box for the same separation task without interheating or intercooling, moving part of the heating duty to lower temperatures and part of the cooling duty to higher temperatures. The same total heat is degraded.

We would like to establish the dimensions for this diagram. We can accomplish this by performing an analysis for the pinch point. Assuming both operating (material balance
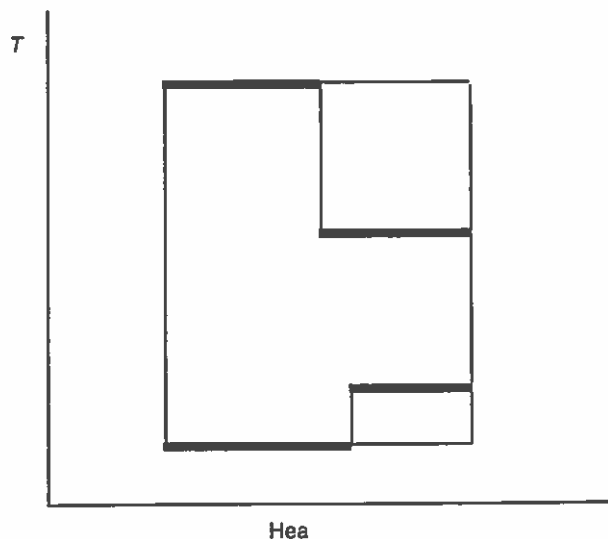
FIGURE 12.6 Expected notched structure for heat cascade diagram for intercooling and interheating.

around top of column) and equilibrium equations ($y_i = \alpha_{ik} x_i / \overline{\alpha}_k$) hold at a pinch and solving for compositions x_i, we get:

$$x_i = \frac{D x_{D,i}}{\dfrac{\alpha_{ik} V}{\overline{\alpha}_k} - L} \tag{12.1}$$

and

$$\sum_i \alpha_{ik} x_i = \overline{\alpha}_k = \sum_i \frac{D x_{D,i}}{\dfrac{\alpha_{ik} V}{\overline{\alpha}_k} - L} \tag{12.2}$$

We proceed as follows, given the flow and composition for the top product.

1. Set reflux ratio R to zero.
2. Compute $L = RD$ and $V = (R + 1)D$.
3. Guess all relative volatilities.
4. Iteratively solve Eq. (12.2) for $\overline{\alpha}_k$. Solve Eq. (12.1) for all x_i.
5. Using a rigorous analysis package, determine the bubble point temperature, $T_{bub}(x_i)$, for this pinch point composition. New (composition, temperature, and pressure dependent) α_i are automatically computed as a part of this calculation.
6. Iterate from step 4 until no changes occur in the variable values. (This computation is rigorous and works even for nonideal physical property behavior.)
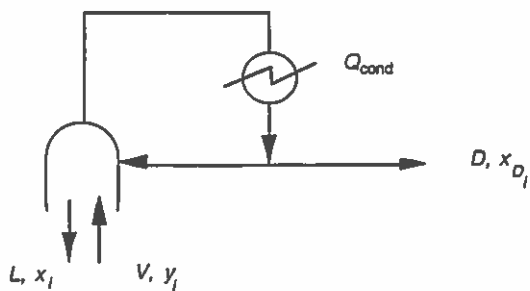
FIGURE 12.7   Top of column.

7. Compute, as follows, Q_{cond} using the heat balance around top of column (see Figure 12.7).

$$Q_{cond} = H_V V - h_L L - h_D D$$

Again, this calculation is also an exact one; no approximations are needed to do it. It will require using a rigorous physical properties package. The values for R and the corresponding values for L and V are *at the pinch point*.

8. Plot the point representing T_{bub} versus Q_{cond} for this value of the reflux ratio, R, on a plot.

9. Increment R by a small amount and repeat until R equals the value required for the normal column.

You will obtain the lower curve shown in Figure 12.8. Note that when the amount of reflux is zero, enough heat must still be removed to condense the top product; thus the heat removal value at the lowest temperature is not zero if the top product is bubble point
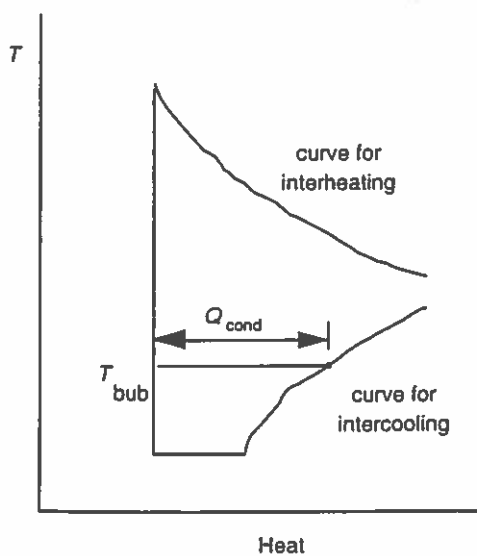

FIGURE 12.8   Intercooling and interheating temperature curves.

liquid. Repeating a similar analysis, stepping the reboil ratio from zero to its value in a normal column, allows one to plot the top curve for interheating.

The bottom curve plotted gives the amount of heat to be removed in the condenser to make the operating line intersect the equilibrium surface at the temperature shown. This is *least* amount of heat that can be removed to get down to this temperature before removing any added heat.

You could remove heat at every stage and keep the steps exactly on the equilibrium surface. The column almost carries out a reversible separation. It fails to be totally reversible (see Fonyó, 1974a, b and Koehler et al., 1992) because the feed is not required to have the same composition as the liquid on the feed tray. The enclosed area for the heating and cooling curves for this case is as small as it can be for the given feed and products; it is a limiting diagram. Of course, one would require an infinite number of stages and infinite area in the exchangers to obtain this performance for the column. Thus, if this limiting diagram is used to formulate heat integration alternatives, one could expect it to yield the best that could be done with the column.

This plot, once completed, allows one to determine the size of the "notches" in the box for the base case that corresponds to intercooling. Figure 12.9 illustrates.

We select a temperature for intercooling, a temperature that is hotter than the condenser temperature. Locate this temperature on the lower curve above and draw a vertical line to the base line shown for the base case (i.e., to the box). We must remove at least the amount of heat to the left of this line from the condenser. We should really remove more so the column does not pinch at the chosen temperature. The amount of heat not removed by the condenser must then be removed in the intercooler.

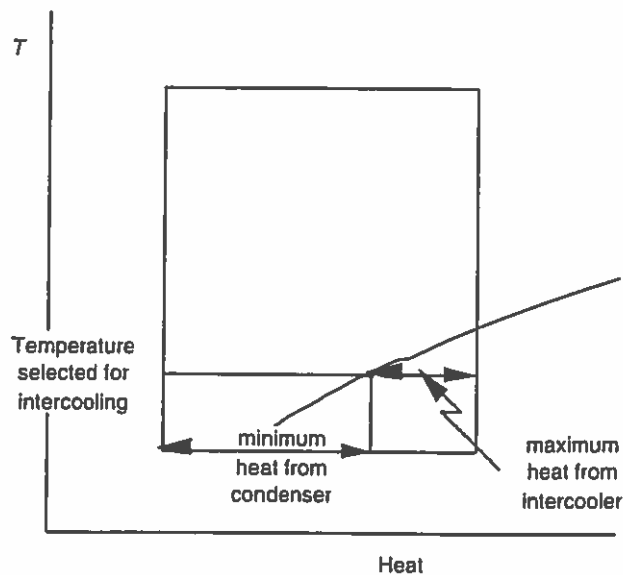A similar construction accounts for interheating.

FIGURE 12.9   Discovering the amount of heat to remove from the condenser and the intercooler.

EFFECT OF CHANGING THERMAL CONDITION OF FEED

We argue here that the curves for intercooling and interheating are valid regardless of the thermal condition of the feed. Examining the method to obtain the intercooling and inter-heating curves, we see that they are a trajectory of pinch points whose T and Q values are determined by stating the top and bottom product compositions and thermal conditions only. Nothing in the analysis involves the thermal condition of the feed; therefore, these curves must be valid whether the feed is a bubble point liquid, dewpoint vapor, two phase, superheated, or subcooled.

We argue that the thermal condition of the feed only changes how far along these curves we proceed before reaching the reflux ratio needed for the top or the corresponding reboil ratio for the bottom. Given the thermal condition of the feed, we can find the reflux and reboil ratios needed by using whatever analysis is appropriate, for example, by using Underwood's method. If the feed is bubble point $(q = 1)$, the heat duties are nearly equal, as argued earlier. If the feed is preheated, the condenser duty will exceed the condenser duty, as shown in Figure 12.10. Here we feed bubble point liquid into a feed preheater that changes its thermal condition. Arguing as before that the sensible heats are small, the heat removed from the condenser, Q_C, has to equal approximately the heat used to preheat the feed, Q_F, plus the heat into the column reboiler, Q_R—that is,

$$Q_C \approx Q_F + Q_R$$

In Figure 12.11, we can parameterize the pinch point curves for determining inter-heating and intercooling with values of q to reflect the thermal condition of the feed into the column. We see that, for the base case of bubble point feed where $q = 1$, we have the box-shaped figure as before. For $q = 0$ (dewpoint vapor), the reboiler heat is less as expected, while the condenser heat *is more* than for the base case. In other words, preheating
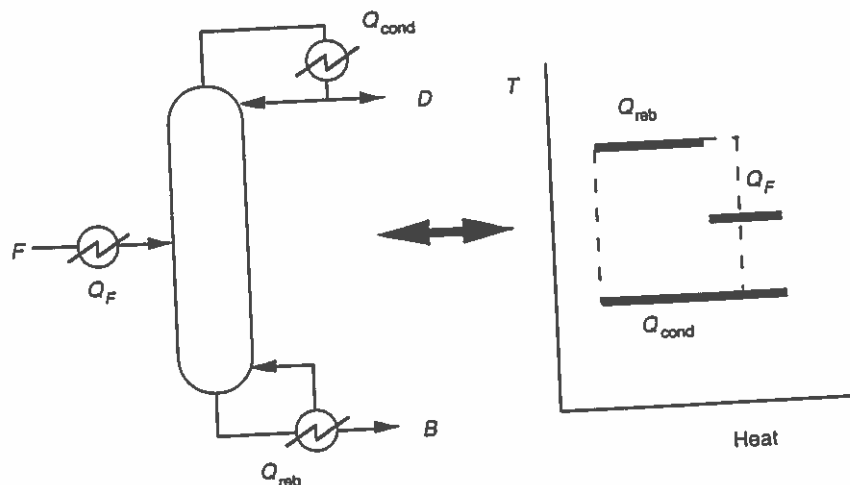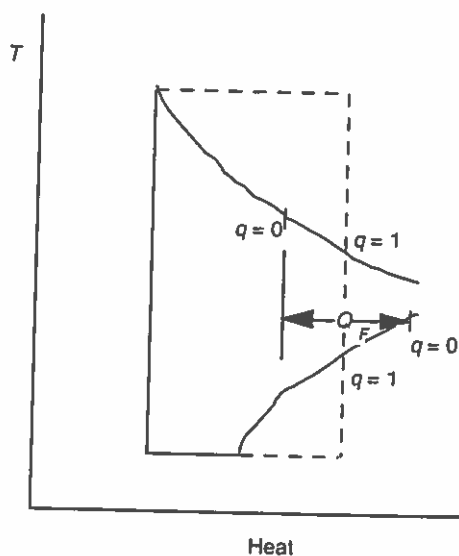

FIGURE 12.10   Preheating the column feed.

FIGURE 12.11   Changing thermal condition of feed. Both duties change when preheating or precooling the feed.

the feed simultaneously *increases* the condenser duty and *decreases* the reboiler duty. By similar arguments, when one precools the feed, the condenser duty reduces and the reboiler duty increases.

The difference in the duties is approximately the amount of heat to change the feed from being a bubble point liquid to the condition being fed to the column.

EXAMPLE FOR USING INTERHEATING/COOLING

Suppose we would like to reduce the utilities required to run two columns that are separating heat-sensitive fatty alcohols. As sketched in Figure 12.12, the column temperatures cannot be increased very much or the alcohols will rapidly decompose in the columns. These temperature limitations preclude the "stacking" of either column on top of the other. We can still get some integration by interheating in one column while intercooling in the other as illustrated in the right-hand side of the figure, carrying out a partial integration for both.

12.1.3   Heat Flows in Side Strippers and Side Enrichers (Carlberg and Westerberg, 1989)

SIDE STRIPPERS

Consider the column configuration shown on the left-hand side of Figure 12.13. This configuration is called a side stripper. As illustrated, such a configuration is capable of separating three ideally behaving species that would normally require the use of two columns. We do see two column shells, each with a reboiler here, but there is only one condenser. We have saved a piece of equipment.
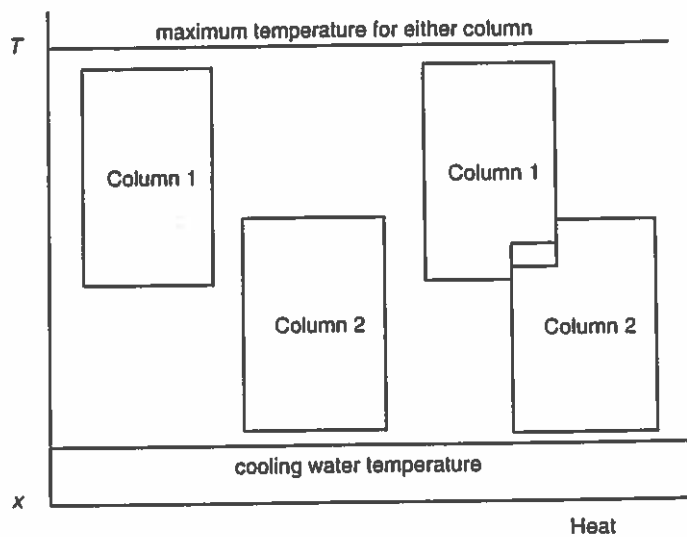
FIGURE 12.12 Example process for which intercooling/interheating is a candidate to improve integration.
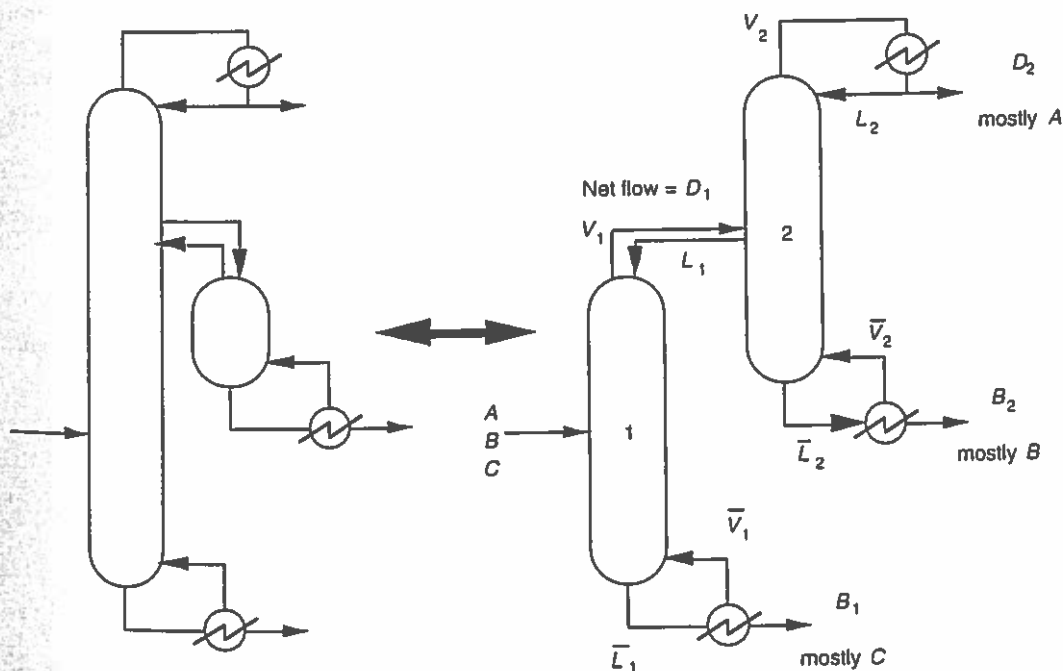

FIGURE 12.13 A side stripper with a topologically equivalent structure to it.

Column simulations show that this configuration requires less heating and cooling than would two separate columns, often as much as 25 to 40% less, so it appears to have a second very interesting advantage. It must have a cost or else it would be more widely used. The disadvantage is, in part, that the temperature drop across it ranges from the bubble point of the top to the bubble point of the bottom, which for this example is from the boiling point of A to the boiling point of C. The top of the column is pressure coupled to the bottom; indeed, since pressure must decrease as one moves up the column (so the vapor will flow up the column), the temperature drop is more than if the column could operate at a fixed pressure throughout (the lowest pressure occurs where the lowest temperature occurs making it even lower relative to the highest). Heat must degrade over this entire range to run this column. With two columns, one can decouple their pressures adjusting them to reduce the temperature drop over which their heat is degraded.

In summary, then, we buy one less exchanger and gain on the first law—often substantially less heat is degraded—but we lose on the second law—it must be degraded ove what is often a much larger temperature drop.

Another point to make for side strippers (and enrichers) is that they are really like hea integrated columns. Therefore, it is inappropriate to consider their use against columns ru using only utilities. We should only compare them from a utility consumption point of vie to columns where we allow the conventional columns to be heat integrated.

Let us first learn how this column performs by developing an approach to analy: it. We shall start by seeing how to compute the minimum reflux for it.

Examination of the second configuration in Figure 12.13 should make it evident is topologically equivalent to the first, but we shall find it is easier to analyze th the first. One can think of the side stripper as two columns. We illustrate it with a component feed—A, B, and C—to make it clearer what the configuration is really doi The first column splits AB from C while the second splits A from B. The side stripper l the separation capability of two columns, but it has only one condenser.

Let us develop the following equations for the second configuration.

$$D_1 = V_1 - L_1$$

We can then write the following:

$$\bar{L}_2 = L_2 + q_2 D_1 = L_2 + q_2(V_1 - L_1)$$

However, we are taking liquid from the second column to provide reflux for the first, ing

$$L_1 = L_2 - \bar{L}_2$$

Solving for q_2 using these two equations, we get

$$q_2 = \frac{-L_1}{V_1 - L_1}$$

We can relate the reflux ratio in the first column to its internal flows, getting

$$R_1 = \frac{L_1}{D_1} = \frac{L_1}{V_1 - L_1}$$

which gives the remarkable result that

$$q_2 = -R_1$$

that is, the thermal condition for the feed to the second column is the negative of the re-flux ratio for the first. Since R_1 is strictly positive, q_2 is strictly negative, which corre-sponds to the net feed to the second column being superheated. One explanation for this is that one is passing vapor to the second column and getting back a part of that vapor as liq-uid. The net flow to the second column can be thought of as the net material flow as vapor plus the heat obtained by cooling the rest from vapor to liquid.

A way to analyze this configuration, then, is the following:

- Establish the bottom and then the top products for the first column.
- Determine the minimum reflux ratio for the first column, using Underwood's method if it is applicable.
- Set the reflux ratio for the first column to some factor (like 1.2) times the minimum reflux ratio for the first column.
- The thermal condition for the feed to the second column is then the negative of this reflux ratio. Determine the minimum reflux ratio for the second column. Set its value to something like $1.2*R_{2,min}$.

SIDE ENRICHERS

The side enricher in Figure 12.14 is also shown in a topologically equivalent form that is easier to analyze.

The analysis here is similar to that for a side enricher. Here we find the thermal con-dition for the feed to the second column is given as:

$$q_2 = \overline{R}_1 + 1$$

that is, it is equal to the reboil ratio for the first column plus one. It will always exceed one, a value that occurs for subcooled liquid feed. The design procedure is precisely the same, except the above should be used to set the thermal condition of the feed for the sec-ond column.

T VERSUS Q DIAGRAMS FOR SIDE STRIPPERS AND ENRICHERS

Let us consider the side stripper just analyzed. We see that it has two reboilers and one condenser. The feed to column 2 is acting like superheated vapor, as we discussed before.

As argued earlier in the section on interheating/cooling, feeding a column with su-perheated vapor simultaneously decreases the reboiler duty, but it also increases the con-denser duty. The heat flows for the side stripper configuration act as if the column T-Q di-agrams have overlapped, as shown in Figure 12.15. The second column reboiler duty has decreased while its condenser duty has increased, consistent with our above observation about preheating its feed.
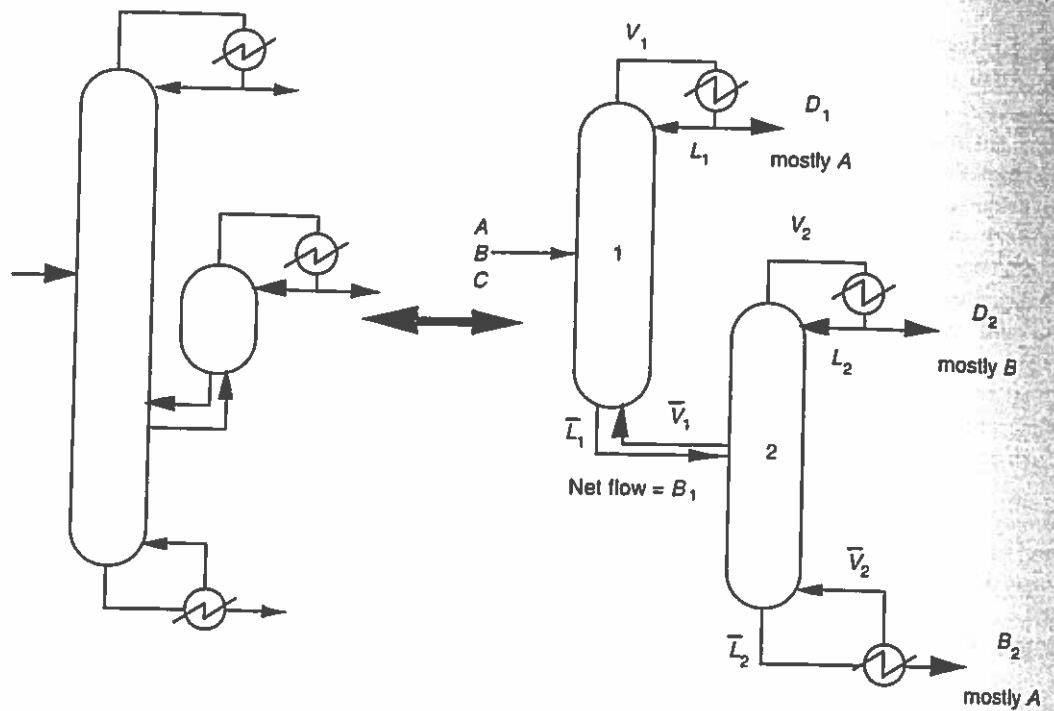
FIGURE 12.14   Side enricher configuration with topologically equivalent structure.
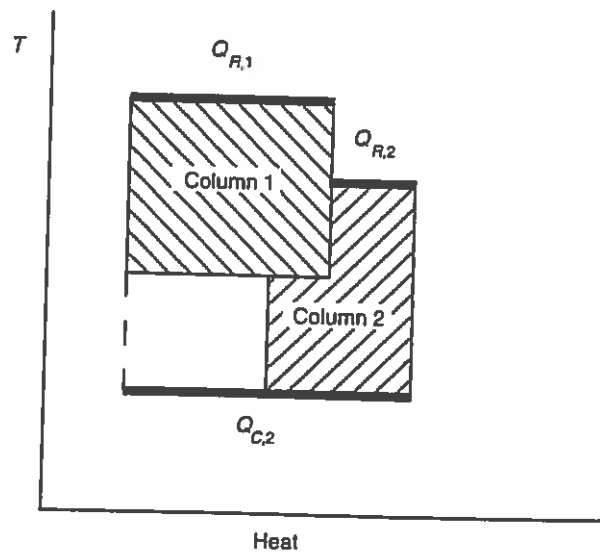

FIGURE 12.15   Heat duties on a T versus heat diagram for side stripper configuration.

This diagram suggests that there could be an advantage to placing a condenser at the top of column 1, allowing its duty to be removed at a higher temperature than from the condenser at the top of the second column.

If the advantage of the higher temperature for heat removal is needed for heat integration, then it may be a good idea. Or, if the extra driving force reduces the heat exchanger area enough, then it may be a good idea.

For a side-enricher configuration, we will simultaneously decrease the condenser duty and increase the reboiler duty, getting a diagram that is the same as the above except that it is flipped vertically, having one reboiler temperature and two condenser temperatures.

For further reading on heat flows in columns, see Dhole and Linnhoff (1992).

REFERENCES

Andrecovich, M. J., & Westerberg, A. W. (1985). A simple synthesis method based on utility bounding for heat-integrated distillation sequences. *AIChEJ.*, **31**, 363.

Carlberg, N., & Westerberg, A. W. (1989). Temperature-heat diagrams for complex columns: 2. Underwood's method for side strippers and enrichers. *I&EC Res.*, **28**, 1379–1386.

Dhole, V. R., & Linnhoff, B. (1992). Distillation column targets. In *Proceedings from the European Symposium on Computer Aided Process Design-I*, 97.

Fonyó, Z. (1974a). Thermodynamic analysis of rectification I. Reversible model of rectification. *Intern. Chem. Eng.*, **14**, 18.

Fonyó, Z., (1974b). Thermodynamic analysis of rectification II. Finite cascade models. *Intern. Chem. Eng.*, **14**, 203.

Koehler, J., Aguirre P., & Blass, E. (1992). Evolutionary thermodynamic synthesis of zeotropic distillation sequences. *Gas Sep. Purif.*, **6**, 4153.

Terranova, B., & Westerberg, A. W. (1989). Temperature-heat diagrams for complex columns: 1. Intercooled/interheated distillation columns. *I&EC Res.*, **28**, 1374–1379.

EXERCISES

A flowsheet simulation program be used to aid in solving the following problems. However, they can also be done using Raoult's law within a spreadsheeting program.

1. Consider a mixture of 35 (mole) % *n*-heptane, 30% *n*-hexane, 10% isobutane, and 25% *n*-pentane. Using Raoult's law, develop the condenser and reboiler heat duties and temperatures for the column separating into two products of two species each for pressures of 1, 5, 10, and 20 atm when running the column at 1.2 times the minimum reflux ratio for it. 99 mole % of the key components should be recovered in their respective products. Use a partial reboiler and a total condenser. The feed to

the column is bubble point liquid. Plot the temperature drop across the column and the average of the two heat duties versus the condenser temperature. Can you notice anything special about this plot? Note, also how close to equal the reboiler and condenser heat duties are for each of the columns (the base case assumption).

2. Repeat the previous exercise but this time do the computation using a commercial flowsheeting system.

3. Repeat the analysis of exercise 1 for the column that produces a distillate that is a single species.

4. Consider the mixture in exercise 1 again. Desired products are all the single component products, each of which is to be 99% pure. Discover the 3-column sequence that requires the least amount of total heat for this separation problem, after the best heat integration you can discover is done between condensers and reboilers. Hot utility is available so heating of a stream up to 425 K is possible; cooling of a stream down to 305 K is possible with cooling water.

5. Repeat the previous exercise, but this time you are allowed to use a maximum of five columns. With five columns, you can propose solutions that involve multi-effecting.

6. You have been asked by another engineer to check over the flowsheet shown in Figure 12.16. Note that the condenser for the first column is a partial condenser. (a) List any obvious design errors. (b) Is the other engineer's analysis believable? (c) Should the feed to column 1 be preheated? Explain your answers.

7. Compute the intercooling/interheating diagram for separating the isobutane from the remaining components at one bar for exercise 1. Assume the feed and products are at their bubble points.
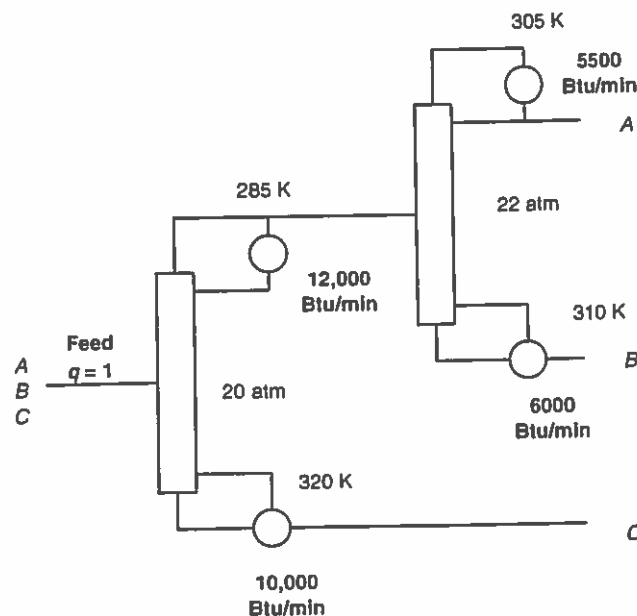

FIGURE 12.16   Separation flowsheet.

8. You are grading a senior chemical engineering design project. The flowsheet the student group proposes contains the separation scheme shown in Figure 12.18. What comments (in red pen) would you make on it? What suggestions would you make for the group to improve·this part of their flowsheet?
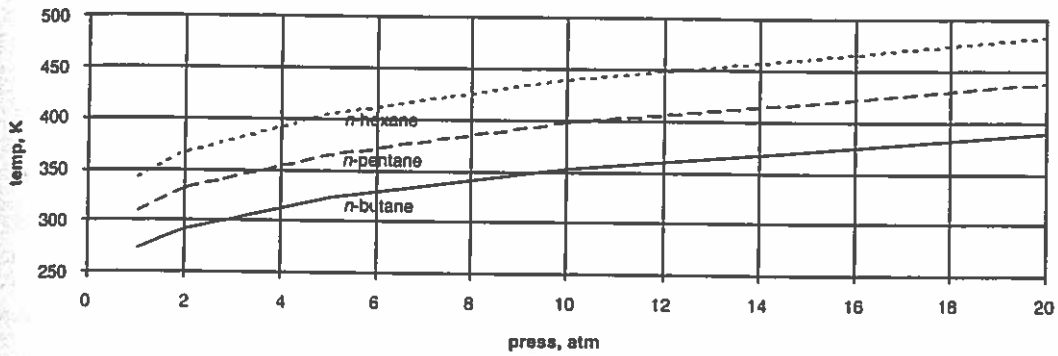

FIGURE 12.17   Temperature vs. vapor pressure for components in Exercise 8.
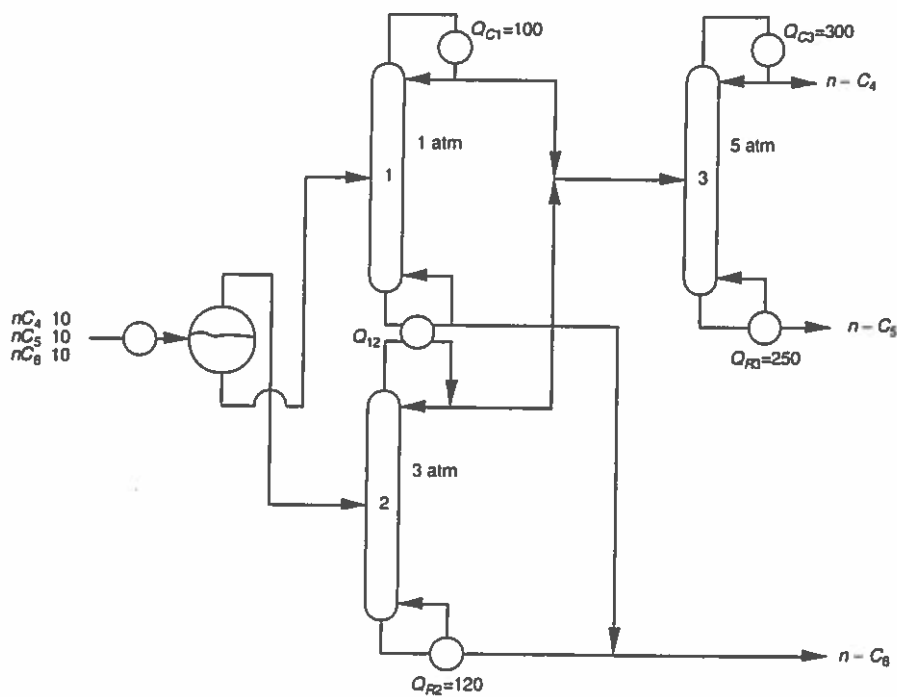

FIGURE 12.18   Proposed separation scheme for Exercise 8.

9. You are given the following mixture. Propose the better heat integrated distillation-based separation sequences to produce five relatively pure single component products. Explain your answer.

Species	Relative Volatility	Amount, kg/s
A		2
	$\alpha_{AB} = 3$	
B		2
	$\alpha_{BC} = 1.15$	
C		2
	$\alpha_{CD} = 3$	
D		1
	$\alpha_{DE} = 1.15$	
E		0.25

10. This problem is a major effort, taking perhaps several tens of hours. Do not casually choose to do it or assign others to do it. Repeat exercise 4—or for the more hearty, exercise 5—but this time worry about the cost of the equipment to carry out the separations and heat exchange. Transfer coefficients for all heat exchangers can be assumed to be 1000 W/m^2 K (assuming both sides are condensing/vaporizing fluids with some fouling having occurred). This problem will require you to consult information not provided here, such as cost estimation correlations for equipment. Remember that the product from a column, if withdrawn as a bubble point liquid and fed to another column, will not be bubble point liquid unless the next column is at the same pressure. To do this problem right, you will have to adjust column pressures to alter the column temperatures and thus reduce the cost for the heat exchangers.