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On the Use of Intermediate Reboilers in the Rectifying Section and Condensers in the Stripping Section of a Distillation Column

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Advantages of an intermediate reboiler in the stripping section of a distillation column and an intermediate condenser in the rectifying section are well-known. For highly nonideal mixtures the distillation process can be improved by placing an intermediate reboiler in the rectifying section of the column or an intermediate condenser in the stripping section, which is counterintuitive. In consequence the more expensive heating utility used in the bottom reboiler can be partially replaced with a less expensive heating medium (at a lower temperature) used in the intermediate reboiler. Similarly, a portion of the condensing duty from the top condenser can be replaced with the less expensive condensing duty (at a higher temperature) in the intermediate condenser. This placement of reboilers (condensers) can be used to reduce the total number of stages in the distillation column with a lower (higher) level of utility. Operating and capital costs of the distillation process with an intermediate heat exchanger can be lower than those for a classic column. A simple method providing valuable insights on the possible placement of intermediate heat exchangers along the column is proposed. It is based on calculations of the vapor flow along the height of a reversible, binary distillation column performing the equivalent separation task.

Introduction

A constant race toward increasing the effectiveness of distillation processes stimulated high interest in a thermodynamically reversible distillation. The reversible binary distillation has been analyzed by Benedict (1947), Flower and Jackson (1964), Fonyo (1974), Fitzmorris and Mah (1980), Koehler et al. (1991), and Dhole and Linhoff (1993). Reversible multicomponent distillation has been examined by Fonyo (1974), Franklin and Wilkinson (1982), and Koehler et al. (1991). The reversible distillation process requires that all the driving forces for momentum, heat, and mass transfer be vanishingly small. In other words, in such a hypothetical process, it is assumed that there is no pressure drop along the column, temperature differences in reboiler(s) and condenser(s) are negligible, and phases coming into contact on a distillation stage (or packing) are at equilibrium. This requires infinitely large heat exchangers and an infinite number of stages (or height of packing) in a distillation column. Note that an infinite number of stages is not a sufficient condition for reversibility of mass transfer (e.g., irreversible column at minimum reflux).

In a classic column with one feed and two products heat is supplied (at a highest temperature) in a reboiler located at the bottom and is removed (at a lowest temperature) in a condenser at the top of the column. In a reversible distillation heat exchange is distributed along the column height. For the stage column model this is realized by an infinite number of intermediate reboilers located in the stripping section (below the feed) and by an infinite number of intermediate condensers in the rectifying section (above the feed). For a differential model of a packed column a continuous heat exchange along the column can be assumed, with a heat supply below the feed and removal above the feed.

Although reversible distillation is not practically attainable, it can serve as a "bench mark" (or "minimum thermodynamic condition"; Dhole and Linhoff, 1993) for measuring the efficiency of a real process (Agrawal and Woodward, 1991). In practical applications, the irreversibilities in a distillation column can significantly be reduced by applying just one intermediate reboiler or intermediate condenser (Terranova and Westerberg, 1989, Fidkowski and Agrawal, 1995). As in any distillation column the corresponding bottom reboiler (or top condenser) duty can be decreased by introduction of an intermediate reboiler (or an intermediate condenser) to the stripping (or rectifying) section of the column. With the intermediate reboiler, a portion of the heating duty is supplied at a temperature lower than the temperature of the bottom reboiler. Similarly, with the intermediate condenser, a portion of the condensing duty is removed at a temperature higher than the temperature of the top condenser. The process becomes more reversible and can be more economical, because of the possibility of using new, less expensive utilities in intermediate heat exchangers.

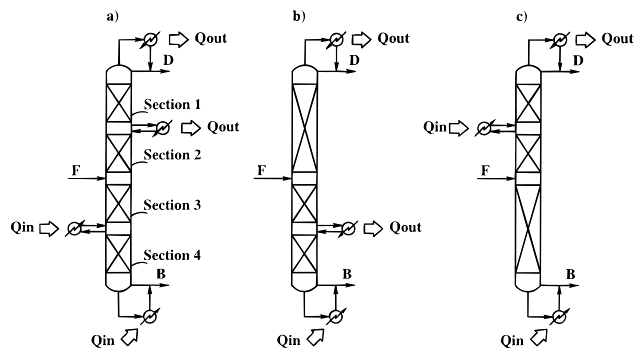

Figure 1. (a) Conventional use of an intermediate reboiler and a condenser in a distillation column. (b) Use of an intermediate condenser in the stripping section. (c) Use of an intermediate reboiler in the rectifying section.

In the known literature the intermediate reboiler is always shown in the stripping section of a column (supplying heat to the column below the feed) and the intermediate condenser is always used in the rectifying section (removing the heat above the feed) (Figure 1a).

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This rule for placement of the intermediate heat exchangers seems to be quite obvious, unquestionable, and widely spread in distillation literature and textbooks (see, for example, King, 1980). On the other hand, Naka et al. (1980) analyzed cases where heat is removed from a distillation column not only in the rectifying section but also in a part of the stripping section (Figure 1b). The objective of this paper is to illustrate these exceptions that exist for nonideal mixtures. We will demonstrate that application of a single intermediate reboiler in a rectifying section for a mixture exhibiting an inflection point on $y-x$ equilibrium diagram may be desirable (Figure 1c). Either we can decrease the bottom reboiler duty (at unchanged reflux ratio) and replace it with a less expensive intermediate reboiler heat or we can decrease the number of stages in the pinch zone without increasing the bottom reboiler duty (in this case the reflux ratio increases correspondingly to the additional heat supplied in the intermediate reboiler). We will also briefly discuss possible savings in the operating and capital costs for a column with such an intermediate heat exchanger.

Intermediate Heat Exchangers in Distillation of an Ideal Mixture

Consider a simple distillation column with one intermediate heat exchanger in the rectifying section and one in the stripping section (Figure 1). We will discuss the possibility of heat exchange in any direction (to and from the column) in both sections. Separation of an ideal mixture is assumed. This assumption will be relaxed later. We will also assume (just for clarity) that the column operates at minimum reflux, i.e., liquid and vapor streams *flowing into* the feed stage are in equilibrium (feed pinch). It is worth mentioning that liquid and vapor streams *leaving* any (theoretical) stage are at equilibrium too. Also, for simplicity, we will illustrate the effects of intermediate heat exchangers, assuming constant molar overflow. This assumption is not necessary and can be omitted by considering additionally enthalpy balance equations.

Equations for operating lines above the feed can be derived from component balances for the upper part of the column, giving:

$$y = \frac{L_i}{V_i}x + \frac{D}{V_i}x_D, \qquad i = 1, 2 \tag{1}$$

The following observations can be made from eq 1:

(1) Both operating lines ($i = 1,2$) have to pass through the distillate composition—point (x_D, x_D) on the $y-x$ diagram

(2) If the intermediate heat exchanger between sections 1 and 2 is a condenser, then $L_1 < L_2$ and $V_1 < V_2$; this information enables us to compare slopes of both operating lines. Rewriting eq 1 for both lines at $x = x_D$, we obtain:

$$\frac{L_1}{V_1} - \frac{L_2}{V_2} = \frac{D}{V_2} - \frac{D}{V_1} < 0 \tag{2}$$

The slope of operating line 1 is lower than the slope of operating line 2. Therefore, operating line 1 is situated above line 2 for $x < x_D$.

(3) If the intermediate heat exchanger between sections 1 and 2 is a reboiler, then and $L_1 > L_2$ and $V_1 > V_2$. The slope of operating line 1 is bigger than the slope of operating line 2, and line 1 is situated below line 2 for $x < x_D$.

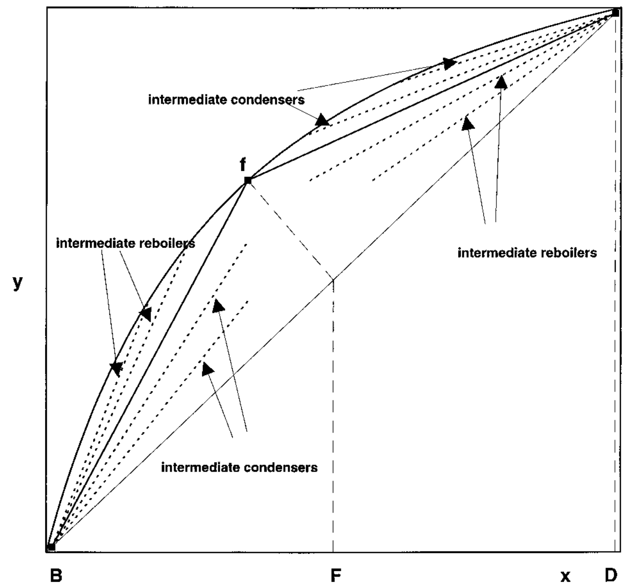

Figure 2. Operating lines for a column with intermediate reboilers or condensers.

Similar conclusions can be derived for operating lines in sections 3 and 4, below the feed. A simple observation can be used to summarize the results: if the flows are larger in any section, due to the introduction of the intermediate heat exchanger, the slope of the operating line in this section will be closer to the diagonal. Figure 2 illustrates possible regions of operating lines when intermediate reboiler or condenser is placed in the rectifying or stripping section.

In a reversible distillation process the phases coming into contact are at equilibrium. Therefore, for a binary mixture, operating lines have to coincide with the equilibrium line. To illustrate how this can be realized, one may approach the reversible process with a column at minimum reflux containing a finite number of reboilers and condensers. The heat exchangers must be placed in such a way so that new operating lines are situated as close to the equilibrium line as possible. According to Figure 2, the intermediate reboilers should be placed in the stripping section and intermediate condensers in the rectifying section. Furthermore, the operating lines for particular sections should be maximally extended up to their intersections with the equilibrium line. Multiple pinch points will occur: the original one, at the feed plate, and one at each heat exchanger location. An example of a $y-x$ diagram for a column with multiple reboilers in the stripping section and multiple condensers in the rectifying section is shown in Figure 3. Except several pinch points, the operating lines do not coincide with the equilibrium line. The reversible process requires an infinite number of heat exchangers along the column.

The approximation of the reversible process, which is shown in Figure 3, helps us notice that the operating lines at the feed stage are parts of the original operating lines for a column without any intermediate heat exchangers. Therefore, the total heating (and cooling) duty in the column with intermediate heat exchangers is the same as in the classic column at minimum reflux. Only the temperatures of the intermediate heat exchangers are different than temperatures of the bottom reboiler and the top condenser. The reversible distil-

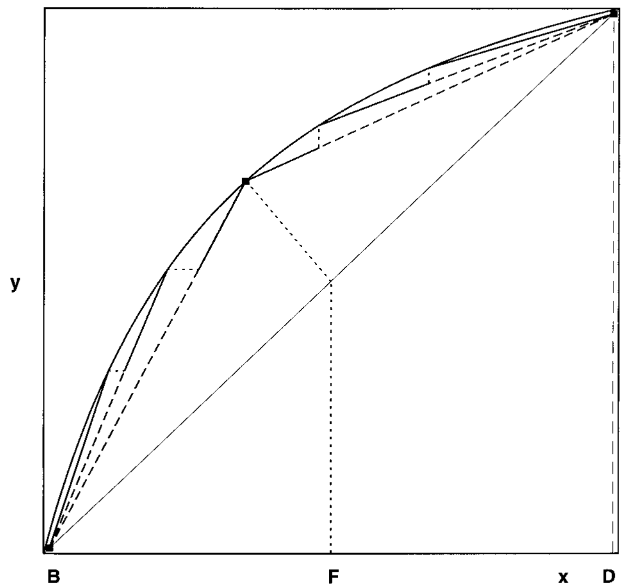

Figure 3. $y-x$ diagram for a column with multiple reboilers in the stripping section and multiple condensers in the rectifying section.

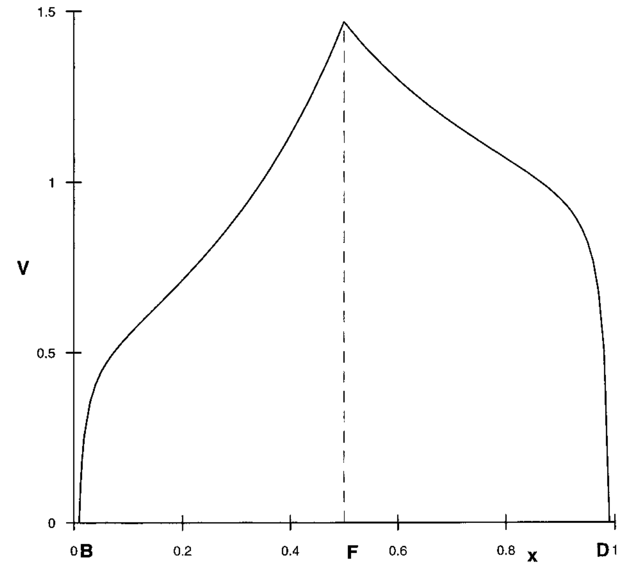

Figure 4. Vapor flow versus composition in the reversible distillation of an ideal mixture for the following data: saturated liquid feed, $F = 1.0$, $x_F = 0.5$, $x_B = 0.01$, $x_D = 0.99$, and $\alpha = 2.0$.

lation, when compared with a classic distillation at a minimum reflux, does not provide any savings in terms of the first law, but it is more attractive in terms of the second law of thermodynamics.

Similarly, it can be shown for multicomponent systems that the energy requirement of the reversible distillation is the same as the energy requirement of a classic column at minimum reflux, with the same product compositions. This fact was also observed by Koehler et al. (1991). Note that only certain product compositions can be achieved in a reversible multicomponent distillation, the main reason being that the feed composition is the same as the composition on the feed stage. This kind of separation (between "direct" and "indirect" split) has been known as a "transition split" (Fidkowski et al., 1993).

The described procedure enabling us to approach the reversible distillation process by placing intermediate heat exchangers along the column height is somewhat intuitive, and it does not provide any quantitative information as to how much duty should be transferred to or from the column at a given height. A more systematic approach is needed to determine the heat duty profile in a reversible distillation. To do that, one can solve material and energy balances for both sections of the column, utilizing also the fact that liquid and vapor phases are at equilibrium at each point. For a constant molar overflow the energy balance equations can be omitted, and instead of the heat distribution simply the vapor flow rate in the column can be calculated. For the stripping section we obtain

$$V = B\frac{x - x_B}{y^*(x) - x} \tag{3}$$

Similarly, for the rectifying section (balancing the lower part of the column)

$$V = \frac{B(x - x_B) + F(x_F - x)}{y^*(x) - x} \tag{4}$$

Application of eqs 3 and 4 for a binary distillation is straightforward. In a reversible distillation of a multicomponent mixture only certain product compositions are feasible. Equations 3 and 4 can also be applied to a multicomponent separation if only these feasible product compositions are used. An example for an ideal binary mixture is shown in Figure 4. We can clearly see that heat energy must be provided to the column below the feed and extracted from the column above the feed. Diagrams like the one on Figure 4 can easily be transformed into cumulative heat duty vs temperature coordinate system (Naka et al., 1980; Koehler et al., 1991) by including additional enthalpy balance equations.

Intermediate Heat Exchangers in Distillation of a Mixture with a Tangent Pinch

Tangent pinches in distillation of a binary mixture correspond to the situation when the operating line becomes tangent to the equilibrium line. An example for a mixture of acetone and water is shown in Figure 5. True minimum reflux is controlled by the value of reflux at which the tangency occurs (continuous operating line in Figure 5), and its value is higher than the minimum reflux calculated assuming pinch conditions at the feed stage (dashed operating line, infeasible). In binary systems this tangency is only possible for the mixtures that exhibit an inflection point on the $y-x$ diagram. Other examples of such mixtures can be found, for example, benzene–ethylenediamine and ethanol–acetic acid. Inflection points on the y-x diagram occur very often for mixtures with azeotropes, like ethanol–water or acetone–methanol. For multicomponent mixtures tangent pinches can be detected as (turning point) singularities in fixed point equations. The examples of binary and multicomponent separations with tangent pinches and a detailed analysis can be found in the works of Levy and Doherty (1986) and Fidkowski et al. (1991).

Mixtures with tangent pinches are frequently difficult to separate—the process is very energy-consuming or requires a lot of theoretical stages. Some hybrid distillation–membrane processes focusing mostly on the reduction of the number of stages in the column with additional power consumption in the membrane module were discussed by Pettersen and Lien (1995) and Stephan et al. (1995).

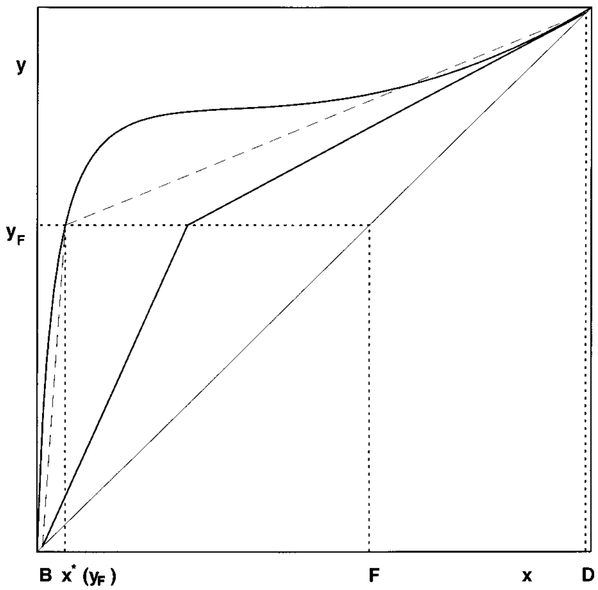

Figure 5. Tangent pinch; example for an acetone−water mixture; $y_F = 0.6$, $x_D = 0.99$, and $x_B = 0.01$. Solid line: feasible operating line tangent to the equilibrium line. Dashed line: infeasible operating line joining distillate composition and feed pinch.

Example 1. Reversible Distillation of a Mixture with a Tangent Pinch. To examine how what the reversible process would look like for a mixture with a tangent pinch, we apply eq 3 and 4 to a separation of a mixture composed of acetone and water in equal amounts to get as a distillate 99 mol % acetone and a bottoms product containing 1% of acetone. The result for the saturated liquid feed is shown in Figure 6a; vapor flow rate in reversible distillation increases from the bottom of the column up to the feed stage and keeps increasing up to a certain point above the feed stage. It means that the heat energy needs to be supplied not only in the stripping section but also along a considerable part of the rectifying section. A result for the saturated vapor feed is shown in Figure 6b. In this figure the difference between the vapor flow above the feed and the vapor feed ($V - F$) is also shown, to see the changes in the vapor flow inside the column due to the intermediate heat exchange only. Note that the vapor flow above the feed (and also the difference $V - F$) decreases, increases, and then decreases again. The reason for this can easily be seen on the $y - x$ diagram (Figure 7), where the operating lines for the rectifying section, in the reversible distillation process, can be drawn from the distillate composition to any point on the equilibrium line at and above the feed pinch. When composition inside the reversible column changes from the feed stage (point G ($x^*(y_F)$, y_F) to the distillate point (x_D, x_D), the slope of this operating line (L/V) first decreases to point A. In this point the operating line becomes tangent to the equilibrium line. It means that directly above the feed heat must be withdrawn from the column to ensure the reversibility of the process. Then, higher in the column, between points A and C, the slope increases (heat must be supplied to the column). At point C the operating and the equilibrium line become tangent again. Finally, between points C and H (distillate) heat is removed from the column again. A similar fact, where the heat duty necessary for the reversible distillation first is supplied in the bottom of the stripping section and then it has to be removed higher in the same (stripping) section, has been noticed by Naka et al. (1980). Koehler et al. (1991) used it to detect

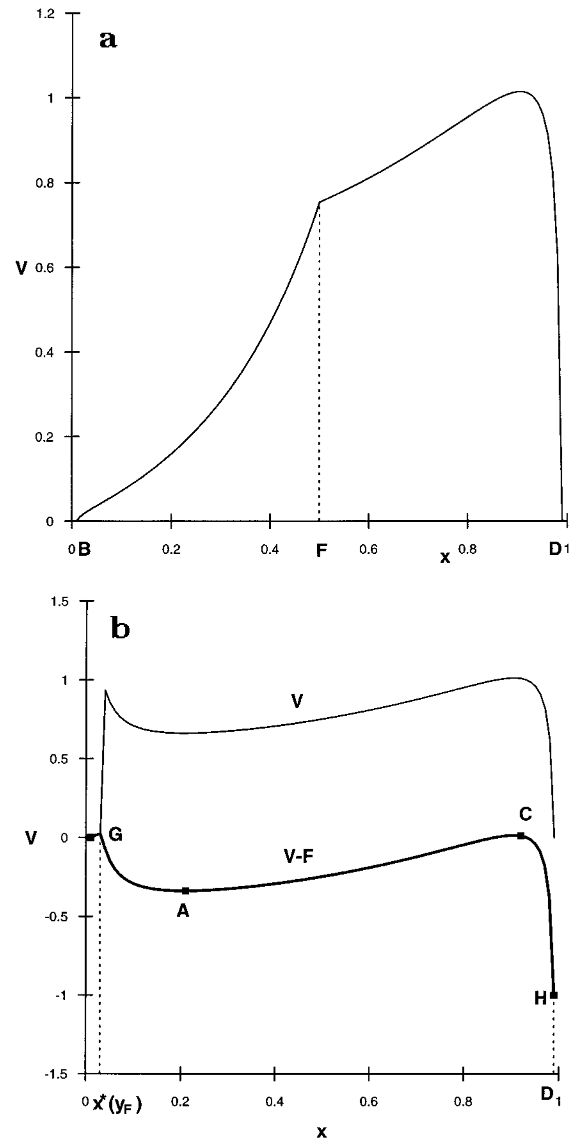

Figure 6. Vapor flow versus composition in the reversible distillation for the mixture of acetone and water, $F = 1.0$, $x_D = 0.99$, and $x_B = 0.01$; example 1. (a) liquid feed, $x_F = 0.5$; (b) vapor feed, $y_F = 0.5$.

tangent pinches in their method of calculating minimum reflux. In the example above the direction of the heat exchange above the feed depends strongly on the feed composition and thermodynamic state. Note that, if our feed ($x_F = 0.5$) is vaporized only partially, so that the feed line points toward point A or somewhere to the right from point A (Figure 7), the need for intermediate condensers directly above the feed vanishes.

A process close to the reversible distillation would be difficult to perform in practice, because of the technical difficulties with supplying certain, strictly determined, portions of heat in the infinite number of differential heat exchangers along the column. Each intermediate heat exchanger brings additional cost and design complications to the process. Another way of realizing an "almost reversible" distillation process is to use a dephlegmator heat exchanger, inside which the separation is also carried out by continuous condensation on the hot side and evaporation of the cold medium (Haselden et al., 1960, Lucadamo et al., 1987). Dephlegmators have at least two disadvantages: the separation effect is not usually high, due to the small interfacial area for the mass transfer. The second problem is that

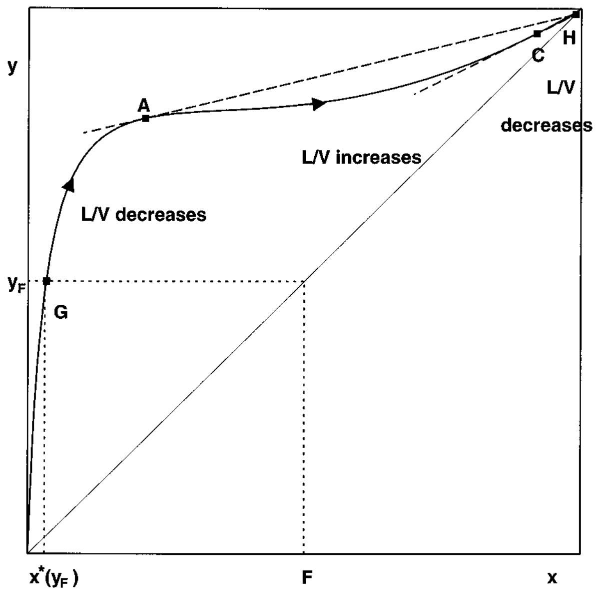

Figure 7. Illustration of changes in the slope of the operating lines in the rectifying section for reversible distillation of an acetone–water mixture (y_F = 0.5, x_D = 0.99).

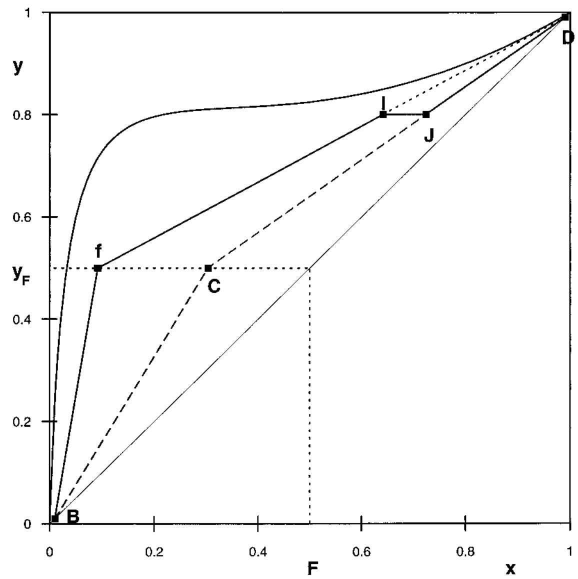

Figure 8. McCabe–Thiele diagram for an acetone–water separation; example 2. BCD: operating line for a column with one reboiler and one condenser. BfIJD: operating line for a column with an intermediate reboiler.

the amount of heat exchanged locally at a certain point of the dephlegmator may differ from the desired heat duty, because it depends only on the state of the hot and the cold stream and it is difficult to control externally.

The observation that heat needs to be supplied in the rectifying section can, however, be used in practice to improve the distillation controlled by a tangent pinch. This improvement may be realized by placing just one intermediate reboiler in the rectifying section of the column having already a bottom reboiler and a top condenser (Figure 1c). The following example illustrates how we can replace a big portion of the expensive bottom reboiler duty with a less expensive intermediate reboiler heat, keeping the total heat duty (and reflux ratio) unchanged.

Example 2. Intermediate Reboiler in the Rectifying Section for a Binary Mixture with a Tangent Pinch To Reduce the Bottom Reboiler Heat Duty. Consider a separation of acetone–water mixture with the following data: P = 101.3 kPa, F = 1, y_F = 0.5, x_D = 0.99, x_B = 0.01 (mole fractions are those of acetone—the more volatile component), and reflux ratio R = 2.5. This problem has been used to teach chemical engineering students the advantage of using a low-temperature waste heat to vaporize the feed (Doherty, 1994). We assume that the feed is already entirely vaporized before entering the column, and we try to further improve this process by placing an intermediate reboiler in the rectifying section of the column. Figure 8 shows a y–x diagram with the equilibrium line and operating line (joining points BCD) for a classic column with one reboiler and one condenser. The reboiler needs to vaporize 0.75 mol of a mixture composed mostly of water (99 mol % water and 1 mol % acetone)/1 mol of feed. In other words, the reboiler duty in terms of vapor flow is V_{reb} = 0.75 at a temperature of 373 K (not including a temperature difference in the reboiler). This corresponds to reflux ratio R = 2.5, which is 1.25 R_{min}. The process requires 36 theoretical stages, and most of the stages are used in the vicinity of the tangent pinch.

We see that at the (approximate) composition range 0.01 < x < 0.80 operating lines for distillation with only one reboiler and one condenser are located far away from the equilibrium line, which is an indication of a high irreversibility of the process. This provides a potential for a further improvement. The operating line for the column with the intermediate reboiler in the rectifying section joins points BfIJD. It has been constructed as follows: knowing the feed and bottoms compositions, we can draw the stripping operating line so that it connects the bottom composition—point B to a point f, located on the line $y = y_F$ and close to the equilibrium line (feed pinch). The choice as to how close the feed stage composition is to the feed pinch is a matter for optimization. Then the operating line for the rectifying section below the intermediate reboiler can be drawn—from point f to the distillate D. This line cannot be used in its entire length, because it either intersects or comes too close to the equilibrium line. Using an intermediate reboiler, we are able to switch from line fD to line CD at point I. Line JD coincides with the original operating line for the rectifying section, when no intermediate reboiler is used (reflux ratio remains unchanged). The location of the intermediate reboiler in the column can also be optimized, since the higher it is situated, the lower is its temperature, but the number of stages in the column increases. In this example we placed it at the height where y = 0.80 and the corresponding temperature is 337 K. The bottom reboiler vapor flow is V_{reb} = 0.10, and the intermediate reboiler generates the remaining vapor necessary for the separation $V_{in.reb}$ = 0.65. The number of stages in this case did not increase considerably—the column requires only 37 theoretical stages (instead of 36). This is because of a big irreversibility of the original process which, although it has significantly improved with the intermediate reboiler, still remains significant. Further improvement can be done by changing the thermodynamic state of the feed. The discussed example deals with a separation of a vapor feed. By vaporizing only a portion of the feed and introducing the liquid and the vapor portion of the feed to the column at different locations (Wankat, 1993, Fidkowski and Agrawal, 1995), a portion of the duty needed previously to vaporize the feed (at 356 K) can be replaced with an even lower level heat in the intermediate reboiler (at about 337 K). We performed also a rigorous simulation of this distillation

Table 1. Results of Calculations for Example 3: Number of Stages as a Function of the Heat Duty in the Intermediate Reboiler

$V_{in.reb.}$	$V_{cond.}$	R	N
0.65	1.75	2.5	37
1.9	3.0	5	21
4.4	5.5	10	18
9.4	10.5	20	17
49.4	50.5	100	16

process, without the constant molar overflow assumption. The results reported above were fully confirmed by the rigorous simulation.

In the example above the two processes (with and without the intermediate reboiler in the rectifying section) are compared at the same total heat duty. In processes where a very high number of stages is required, one may find it useful to increase the intermediate reboiler duty in order to relieve locally the tangent pinch and decrease the total number of stages. It decreases the reversibility of the process (locally, in the vicinity of the tangent pinch), but it may still be profitable if the low level heat used in the intermediate reboiler and the cooling medium are inexpensive. Example 3 illustrates this application.

Example 3. Intermediate Reboiler in the Rectifying Section for a Binary Mixture with a Tangent Pinch To Reduce the Number of Stages. For the same data as in example 2 we examine the effect of additional heat duty in the intermediate reboiler on the number of stages. The minimum number of theoretical stages to separate an acetone−water mixture to obtain $x_D = 0.99$ and $x_B = 0.01$ is 14 and we had 37 stages for the intermediate reboiler case above, so there is a potential to reduce the number of stages in the column. The reduction in the number of stages is achieved by increasing the intermediate reboiler duty and condenser duty (reflux ratio), correspondingly. The bottom reboiler duty remains unchanged in all calculations ($V_{reb} = 0.10$). The results are shown in Table 1. We are able to achieve a significant reduction in column height at the expense of increased consumption of intermediate reboiler and condenser utilities and also at the expense of the increased column diameter in the top section. This may be especially useful for very tall columns, and further economic calculations are needed to evaluate the profitability of such a tradeoff.

In many practical cases it may be difficult to find a low level heat source. Then the concept of a heat pump can be used together with the intermediate reboiler. More information on heat pump applications in distillation, with and without intermediate heat exchangers, can be found in papers by Null (1976) and Agrawal and Yee (1994). A small temperature difference between the top of the distillation column and the intermediate reboiler location in the rectifying section could make the heat pump concept viable for these applications.

Tangent pinches are also common in mixtures containing more than two components. We applied the concept of the intermediate reboiler in the rectifying section for a ternary mixture of acetaldehyde, methanol, and water. Analysis of a tangent pinch behavior for this mixture was described by Fidkowski et al. (1991). Simulation results for this mixture indicated that the intermediate reboiler had to be located close to the feed stage in order to place it below the tangent pinch. Therefore, the intermediate reboiler temperature was almost the same as the feed stage temperature, and it would be much more convenient in this case to vaporize the feed than design a column with the intermediate reboiler. For the vapor feed and for the considered feed and product compositions, the tangent pinch no longer controls the separation of this mixture. It was shown, however, that it is possible to replace a portion of the bottom reboiler duty with the intermediate reboiler located in the rectifying section.

Conclusions

In the examples above we demonstrated that the reversibility of the distillation process for a nonideal mixture can be substantially increased by placing an intermediate reboiler in the rectifying section. Certainly this situation can be symmetrically reversed, and an intermediate condenser in the stripping section can be applied in the case when the tangent pinch is located below the feed. Note that this is against the widely accepted, normal distillation practice.

What finally decides about a practical application of the process is its total cost consisting of the operating cost and capital cost. Operating cost comparison depends on its availability and prices of heating media at various temperatures. In example 2 we can replace 87% of high-temperature heat duty (provided at 373 K in the bottom reboiler) with a lower level heat duty in the intermediate reboiler (at 337 K). We can furthermore replace also a significant portion of the heating medium used to vaporize the feed (at 356 K) with the intermediate reboiler heat. Additional heat duty supplied to the intermediate reboiler enables us to decrease significantly the number of stages in the column (example 3).

The lower level heat duty might be available as a “waste heat” that is free or inexpensive. Therefore, there is a potential for significant operating cost savings. Usually there is a tradeoff between the capital and energy costs. Certainly the column with an intermediate heat exchanger is more complex than the classic column, which suggests that its capital cost might be higher. On the other hand, both columns have almost the same number of stages. Furthermore, the total heat duty is the same in both cases, which means that the total heat exchanger area will not change (with the same temperature difference in heat exchangers). In a situation when a low-temperature heat duty is not available, one may decide to use a heat pump. Another solution is to provide a portion of the heat duty to the bottom reboiler and another portion of it to the intermediate reboiler at one high temperature. Then the total area of heat exchangers (the bottom reboiler and the intermediate reboiler) will be smaller than the area of the bottom reboiler in the classic column, due to a higher temperature difference in the intermediate reboiler. Last, the diameter of the column below the intermediate reboiler is much lower than the diameter of the classic column, due to the decreased traffic of fluids. In summary, a potential exists to improve the operating cost and capital cost of a distillation column by using an intermediate reboiler below the pinch (may include a portion of the rectifying section above the feed) and an intermediate condenser above the pinch (may include a portion of the stripping section below the feed).

Acknowledgment

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Notation

B = bottom product flow rate, mol/s
D = distillate flow rate, mol/s
F = feed flow rate, mol/s
L = liquid flow rate, mol/s
N = number of theoretical stages
R = reflux ratio
V = vapor flow rate, mol/s
x = mole fraction of the more volatile component in the liquid phase
y = mole fraction of the more volatile component in the vapor phase
α = relative volatility

Subscripts

1, 2, 3, 4 = column section number
i = column section
B = bottoms
cond. = condenser
D = distillate
F = feed
f = feed stage
reb. = reboiler
in.reb. = intermediate reboiler

Superscripts

* = equilibrium

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