Prefractionation to Reduce Energy Consumption in Distillation without Changing Utility Temperatures

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Distillation systems composed of a prefractionator and a main column are introduced. The columns operate at different pressures, but the temperatures of the utilities are the same as for a single column performing the same separation. The energy integration of the columns is realized by condensing a vapor overhead from the prefractionator in an intermediate reboiler located in the stripping section of the main column or by boiling a bottoms liquid from the prefractionator in an intermediate condenser placed in the rectifying section of the main column.

Analysis is performed to determine feasible regions of operational parameters and to minimize vapor flows. A detailed simulation example of the benzene–toluene separation shows that for certain feed compositions it is possible to reduce energy requirements of a conventional binary distillation column by almost one-third using this new distillation system. This savings is achieved without any change in the temperature of the utilities used.

Introduction

Although distillation is one of the oldest and best known unit operations, the subject of energy savings in distillation is continuously discussed in literature. A distillation column can be represented as a device that consumes heat at a higher temperature (in a reboiler) and rejects heat at a lower temperature (in a condenser) to produce separation work. If the degraded heat rejected from the condenser of one column can be reused in another distillation column, then the separation in this second column can be performed without consuming any additional external energy. Some early examples of heat-integrated distillation systems originated in air separation, where so-called double-column arrangement has been known since 1905 (Thevenot, 1979).

The number of possible heat-integrated distillation systems is constrained by the second law of thermodynamics, since the temperature in the condenser rejecting the heat must be high enough to ensure heat transfer to the reboiler of the column that reuses this heat. Reboiler and condenser temperatures can be matched by adjusting the pressures in the distillation columns. The whole synthesis task consists of a series of distillation synthesis problems (for various pressures) combined with synthesis of appropriate heat-exchanger networks.

Since the energy is finally rejected back from the distillation system to the cooling utility, the term *energy consumption* (or *energy savings*) is not precise and should rather be replaced by *utility consumption* or *exergy consumption.* On the other hand, the degraded energy in the cooling medium is usually lost, which justifies the usage of the *energy consumption* term. We will use the terms *energy consumption* (savings) and *utility consumption* (savings) interchangeably.

Methods for the synthesis of multicomponent separation systems were initially formulated as computationally expensive mathematical tasks, for example, a dynamic programming-based algorithm (Rathore et al., 1974) or the branch-and-bound method (Sophos et al., 1978; Faith and Morari, 1979). A general evolutionary technique for synthesizing an energy-efficient process was proposed by Umeda et al. (1979). This is a graphical approach, where heating and cooling curves of available heat sources and heat sinks are plotted on a heat-availability diagram, and the process undergoes evolutionary changes in order to minimize utility requirements and relieve possible "pinch points," or the points where heat transfer may be difficult or impossible due to a vanishingly small temperature differences. This type of "pinch" analysis has been widely developed by various authors (Smith and Linhoff, 1988; Linhoff and Dhole, 1992; Naka et al., 1982).

Andrecovich and Westerberg (1983) proposed a graphical

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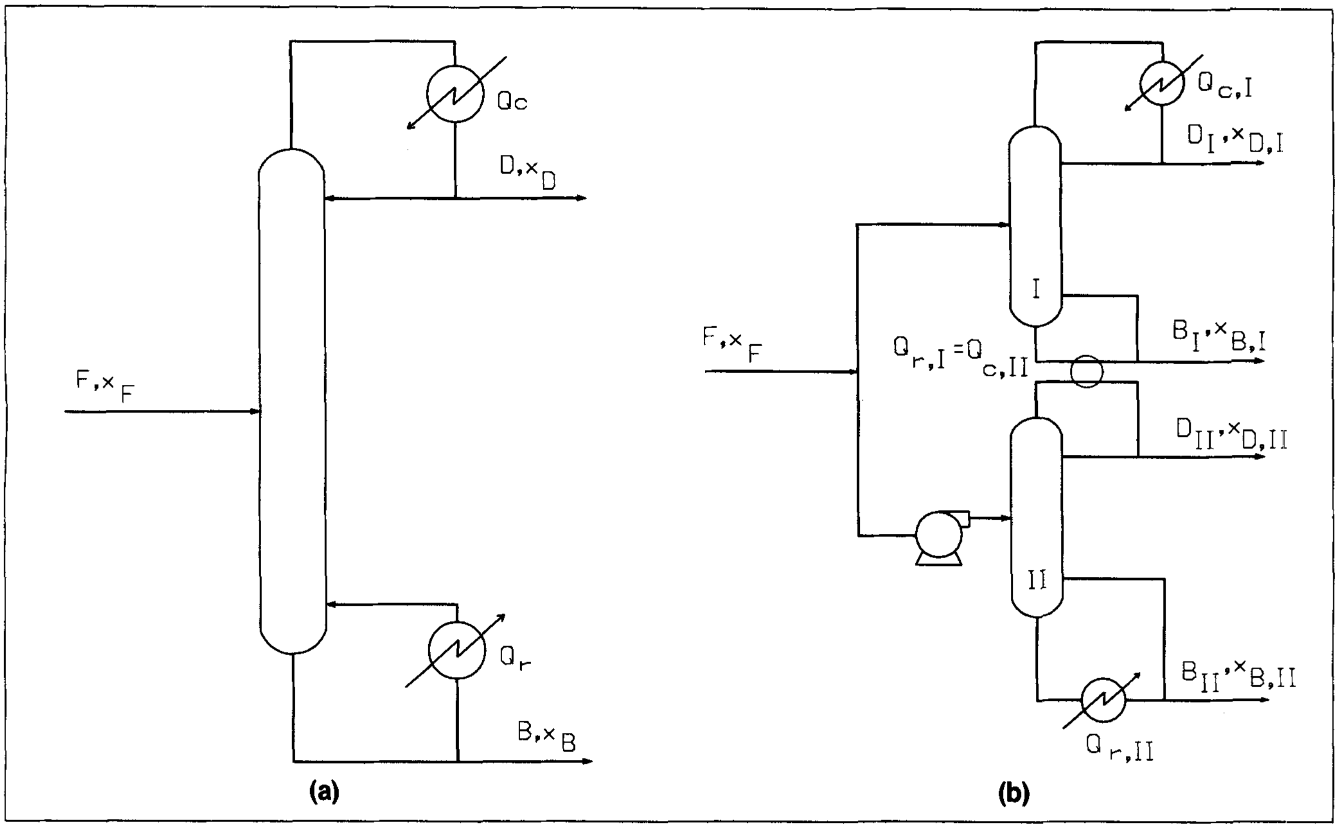


Figure 1. (a) Single-distillation column; (b) heat-integrated system of two columns for multieffect distillation.

representation of distillation processes on a temperature–heat duty (T-Q) diagram, similar to the one used in the heat-exchanger synthesis procedures. It proved to be very useful in the synthesis of separation systems since it enables a quick quantitative comparison of various distillation systems (Westerberg, 1985; Terranova and Westerberg, 1989; Carlberg and Westerberg, 1989a,b).

Another type of column integration for multicomponent systems is realized in so-called complex columns (Petlyuk et al., 1965; Carlberg and Westerberg, 1989a,b; Agrawal, 1993), where energy savings are achieved by a proper arrangement of material and/or heat flows between distillation columns with multiple feeds and side streams.

Heat integration can also be applied to binary distillation (so-called "multieffect distillation"—Wankat, 1993), and the obtained column configurations can then be included in multicomponent separation flowsheets.

For binary separation, heat-integrated distillation systems that have lower utility consumption than nonintegrated systems require that the temperature difference between the heating and cooling media be increased. Consequently, either a warmer heating medium or a colder cooling utility must be used for the integrated system. For example, a well-known multieffect distillation process, shown in Figure 1b (Robinson and Gilliland, 1950; Wankat, 1993), can be compared with a separation process performed in a single column (Figure 1a). Comparison of these cases is very clear on the T-Q diagram shown in Figure 2: a single column consumes approximately twice as much heat as the double-column system, but the heat is provided to the single column at a much lower temperature. Clearly, in order to reduce heat input, "first law heat" is traded for "second law ΔT's," and one does not "gain something for nothing" (Westerberg, 1985). This aspect of heat-integrated systems decreases their attractiveness, because warming utilities with increased temperatures (and cooling media with lower temperatures) are more expensive and sometimes (especially in cryogenic systems) these cost differences could be substantial.

Adding an intermediate reboiler to a column (Terranova and Westerberg, 1989; Fidkowski and Agrawal, 1995), does not decrease total heat input requirements, but it does allow

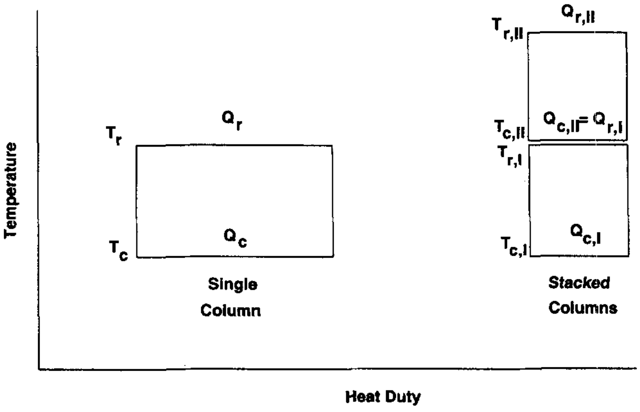


Figure 2. Temperature–heat duty diagram for a single column and for the multieffect distillation process.

a portion of the heat to be provided at a lower temperature than for a classic column with one reboiler. Similarly, the use of an intermediate condenser does not alter the total heat requirement of the distillation, but allows some of this heat to be rejected at a higher temperature.

A combination of an intermediate reboiler or intermediate condenser together with energy integration of distillation columns operating at different pressures may lead to new column sequences, where first law heat is no longer traded for the second law temperature difference. The objective of the present work is to propose a type of heat integration in distillation where the utility temperatures remain the same as for nonintegrated systems, but system energy requirements are reduced. We do not claim gaining something for nothing. The savings are achieved by changing the distillation equipment rather than the utilities.

One example of such a distillation system has recently been proposed independently by Rathbone (1995) as part of a larger distillation configuration for air separation. Some other possibilities of energy-integrated distillation systems that meet the criterion of unchanged temperature of utilities are discussed below.

Prefractionator Concept: Optimization of Minimum Vapor Flow

We will discuss the separation of *binary* mixtures in a single distillation column and in a proposed prefractionator system comprising columns integrated by *heat and mass* streams. A detailed analysis is first presented for comparing one of the proposed prefractionator schemes with a single distillation column. It is followed by a brief discussion of other prefractionator schemes. In the analysis, energy requirements of separation at minimum reflux in a prefractionator system and in a single column will be compared. For simplicity, constant molar overflow in the columns is assumed. Changes of heat of vaporization with pressure will also be neglected at the discussed pressure range. Feed is introduced to the separation system (i.e., either to the single column or to the prefractionator system) as a saturated liquid.

Binary distillation in a single column

The classic distillation column is shown in Figure 1a. A boiling mixture of composition, x_F (mole fraction of the more volatile component), is separated in a single distillation column into a distillate of composition, x_D, and a bottoms product of composition, x_B, both withdrawn as saturated liquids. This information, together with the feed flow rate, is sufficient to calculate distillate (and bottoms) flows. From the overall material balance and component balance we obtain

$$D = F\frac{x_F - x_B}{x_D - x_B}. \qquad (1)$$

Minimum vapor flow is calculated from the material balance of the rectifying section (or stripping section) and equilibrium condition at the feed level, which gives

$$V_0^M = D\frac{x_D - x_F}{y_F - x_F}. \qquad (2)$$

Column with a prefractionator

Let us consider the same separation task performed in a distillation system shown in Figure 3, consisting of a prefractionator (I) and a main column (II). We assume that the entire feed stream is introduced to the first column, although in general this feed stream could be split into two fractions, each one entering a separate column. The prefractionator operates at a higher pressure than the main column. It produces only a portion of the distillate (D_{I}) at the desired composition x_D. The bottoms product from the prefractionator (B_{I}), having composition $x_{B,I}$, is introduced as a feed to the main column. The other portion of the distillate (D_{II}) and the bottoms product (B) is produced from the main column. The heat withdrawn from the condenser of the prefractionator is passed to the intermediate reboiler located in the main column below the feed to this column. Generally the composition of two distillate streams D_{I} and D_{II} is the same. Therefore, a rectification section must be used in the prefractionating column. The use of a stripping section in the prefractionator is optional.

In a general case, the prefractionator can be heated with a heating medium that is either at a lower or higher temperature than the heating medium for the bottom reboiler in the main column. However, the case of interest for this article is where the bottoms of both columns are at the same temperature. The pressure in the prefractionator is therefore set so that the temperatures at the bottoms of the two columns are the same. This means that the same heating medium can be used in the prefractionator reboiler and in the bottom reboiler of the main column (for similar reboiler ΔTs). Since the bottom product leaving the prefractionator is richer in the more volatile component than the final bottom product in the main column, the pressure in the prefractionator column must be higher than in the main column. The pressure in the main column is set to be the same as in the case of distillation in a single column; therefore, the same cooling medium can be used for the single column and for the column with the prefractionator.

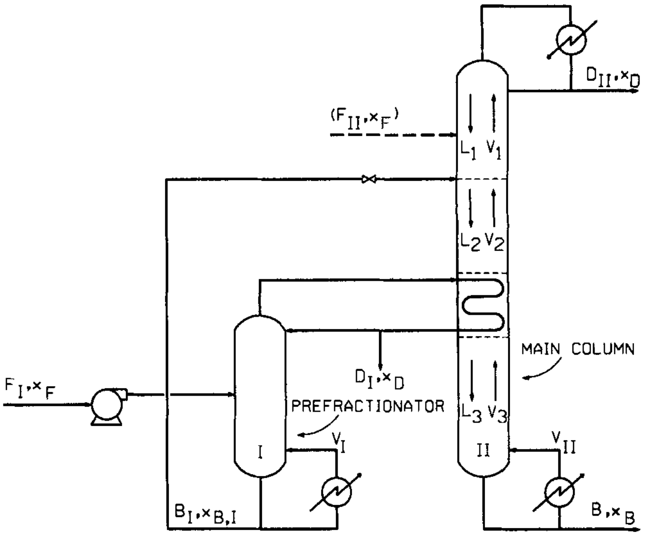


Figure 3. Prefractionator configuration with an intermediate reboiler.

The objective of this analysis is to show that at minimum reflux conditions, and for certain values of process parameters, the energy of separation in the prefractionator system is lower than the energy required for a single column.

We will assume that (for our mixture at given operating conditions) the heat transfer in the intermediate reboiler is possible in the desired direction: from the prefractionator to the main column. We also assume that the feeds to both columns are saturated liquids. More accurate calculations would show that, because of the pressure changes, the feed to the prefractionator column is slightly subcooled and the feed to the main column contains a certain amount of vapor. These small effects will be neglected in the following approximate analysis.

Let us define d as the recovery of the distillate from the prefractionator ($0 < d < 1$), as follows:

$$d = \frac{D_I}{D}, \tag{3}$$

where

$$D = D_I + D_{II}. \tag{4}$$

The bottoms flow rate and composition in the prefractionator can be calculated from material balance and expressed in terms of d as

$$B_I = F - dD \tag{5}$$

$$x_{B,I} = \frac{Fx_F - dDx_D}{F - dD} \tag{6}$$

or, after substituting Eq. 1 for D,

$$x_{B,I} = \frac{x_F(x_D - x_B) - dx_D(x_F - x_B)}{(x_D - x_B) - d(x_F - x_B)}. \tag{7}$$

Minimum vapor flow in the first column is calculated from the equation analogous to Eq. 2

$$V_I^M = D_I \frac{x_D - x_F}{y_F - x_F}. \tag{8}$$

Internal vapor flow rates in the main column (Figure 3) are related as follows:

$$V_2 = V_1 \tag{9}$$

$$V_3 = V_2 - V_I, \tag{10}$$

and at minimum reflux

$$V_{II}^M = V_3^M. \tag{11}$$

The total minimum vapor flow for the prefractionator and the main column is equal to the sum of the minimum vapor flows for these columns and is also equal to the vapor flow rate in the intermediate and top sections of the main column:

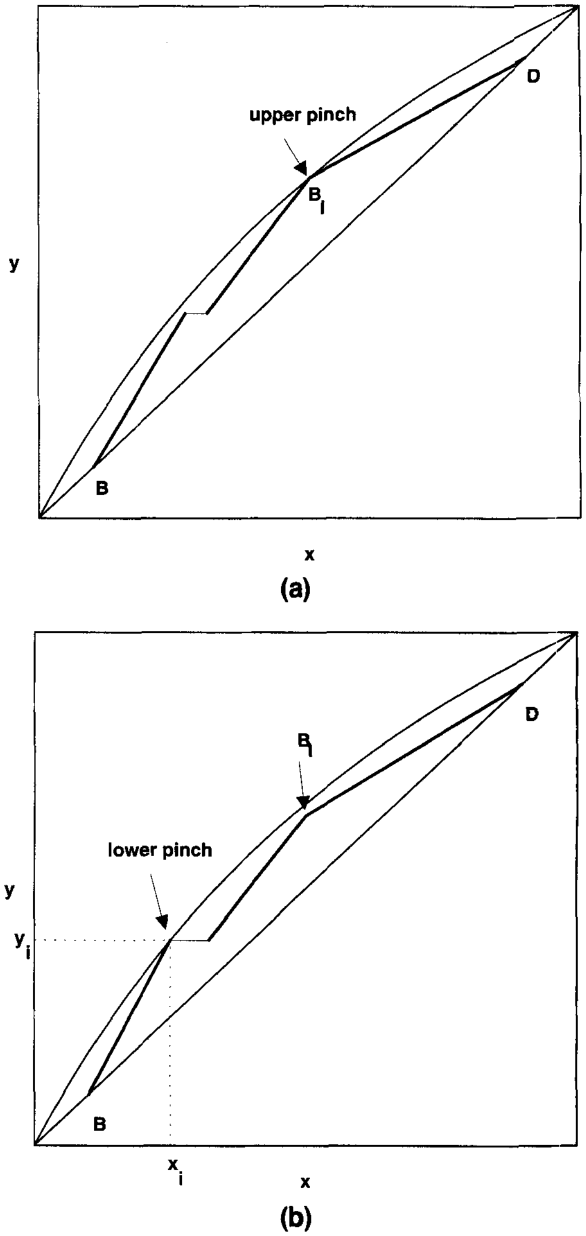


Figure 4. Operating lines on y-x diagram for the main column in the prefractionator system with an intermediate reboiler, at minimum reflux conditions.

(a) Upper pinch controls; (b) lower pinch controls.

$$V_P^M = V_I^M + V_{II}^M = V_2^M \tag{12}$$

The main column operates under minimum reflux conditions when the operating line coincides with the equilibrium line. The pinch may occur at the feed level (Figure 4a), at the intermediate reboiler level (Figure 4b), or in both places at once (Fidkowski and Agrawal, 1995). We will denote the minimum vapor flow in the intermediate section of the main column corresponding to the feed (*upper*) pinch as V_2^{up}, and the one corresponding to the intermediate reboiler (*lower*) pinch as V_2^{lo}. The controlling pinch sets the value of minimum vapor flow for the column system

$$V_P^M = \max\{V_2^{\mathrm{lo}}, V_2^{\mathrm{up}}\}. \tag{13}$$

The minimum vapor flow for the case when the feed pinch controls is calculated from the material balance of the rectifying section of the main column, assuming equilibrium between liquid and vapor phases at the feed level, resulting in

$$V_1 = D_{\mathrm{II}} \frac{x_D - x_{B,\mathrm{I}}}{y_{B,\mathrm{I}} - x_{B,\mathrm{I}}}. \tag{14}$$

From Eqs. 3, 4, 9 and 14 we have

$$V_2^{\mathrm{up}} = (1-d) D \frac{x_D - x_{B,\mathrm{I}}}{y_{B,\mathrm{I}} - x_{B,\mathrm{I}}}, \tag{15}$$

where $x_{B,\mathrm{I}}$ is given by Eq. 7.

In the case when the lower pinch controls, a material balance for the stripping section of the main column is

$$L_3 = V_3 + B \tag{16}$$

$$L_3 x_i = V_3 y_i + B x_B, \tag{17}$$

where x_i, y_i are the compositions of the liquid and vapor phases, in equilibrium, at the intermediate reboiler level. The minimum vapor flow in this section can be calculated from Eqs. 16 and 17 as

$$V_3 = B \frac{x_i - x_B}{y_i - x_i}. \tag{18}$$

The minimum vapor flow in the intermediate section is then calculated from Eqs. 8, 10, and 18

$$V_2^{\mathrm{lo}} = dD \frac{x_D - x_F}{y_F - x_F} + B \frac{x_i - x_B}{y_i - x_i}. \tag{19}$$

From Eqs. 13, 15, and 19 we obtain a more explicit form for the expression for minimum vapor flow in a distillation system with prefractionator:

$$V_P^M = \max\left\{ dD \frac{x_D - x_F}{y_F - x_F} + B \frac{x_i - x_B}{y_i - x_i}, \right. \\ \left. (1-d) D \frac{x_D - x_{B,\mathrm{I}}}{y_{B,\mathrm{I}} - x_{B,\mathrm{I}}} \right\}. \tag{20}$$

To compare the minimum energy requirements of the prefractionator column system with those for a single column, we calculate the ratio of minimum vapor flows from Eqs. 20 and 2

$$\frac{V_P^M}{V_0^M} = \max\left\{ d + \frac{y_F - x_F}{x_F - x_B} \frac{x_i - x_B}{y_i - x_i}, \right. \\ \left. (1-d) \frac{y_F - x_F}{x_D - x_F} \frac{x_D - x_{B,\mathrm{I}}}{y_{B,\mathrm{I}} - x_{B,\mathrm{I}}} \right\}. \tag{21}$$

For given compositions of feed, distillate, and bottoms, the ratio of minimum vapor flows depends on the distillate recovery in the prefractionator column, d, and on the liquid composition in the main column at the intermediate reboiler level, x_i. The distillate recovery from the prefractionator is a manipulated variable and can be set during the operation, independently of any process data. The composition of the liquid vaporized by the intermediate reboiler depends on all the process parameters and on the location of the intermediate reboiler in the column. The intermediate reboiler should be located in the main column at the lowest possible level where the temperature difference between condensing overhead vapor from the prefractionator and the liquid reboiled in the main column is sufficient for satisfactory heat transfer. Therefore the value of x_i depends on the actual pressures in the prefractionator column and in the main column, the distillate composition, the physical properties of the mixture (namely on the bubble and dew points of the mixture and how they change with composition and pressure), and on the heat-transfer area of the intermediate reboiler. All these factors could be included in our screening analysis, but the problem is losing its general formulation and becoming very case-specific. Instead, we will perform our analysis for several feasible values of x_i and examine the spectrum of results obtained to see whether there is any potential for energy savings.

Let us consider the example of the separation of an ideal binary mixture with the following data: $x_F = 0.8$, $x_D = 0.99$, $x_B = 0.04$, $\alpha_{\mathrm{I}} = 1.9$, $\alpha_{\mathrm{II}} = 2.0$. Notice that relative volatility in the prefractionator column is lower than in the second (main) column to account for different pressures. The ratio of minimum vapor flows was calculated from Eq. 21 (with $x_{B,\mathrm{I}}$ given by Eq. 7) and is shown in Figure 5 as a function of distillate recovery in the first (prefractionator) column, for various x_i ($x_B < x_i < x_{B,\mathrm{I}}$). We can see that the ratio of minimum vapor

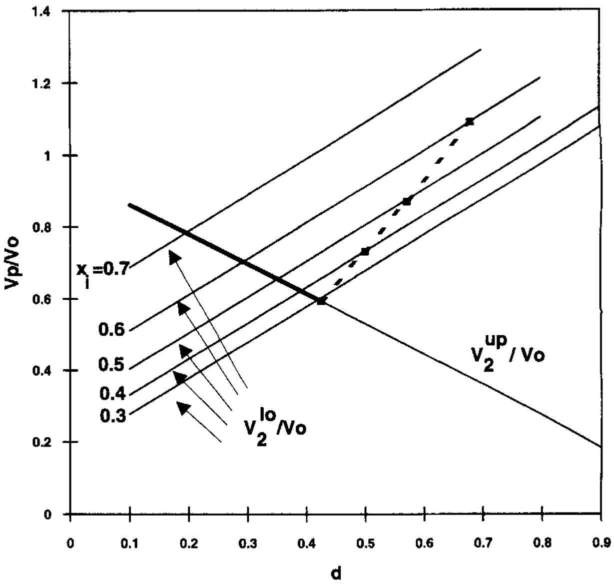


Figure 5. Ratio of minimum vapor flow in the prefractionator system and single column as a function of distillate recovery from the prefractionator.

flows given by Eq. 21 exhibits a minimum for any fixed value of x_i. The left side of the curve (with negative slope) is V_2^{up}/V_0. For small values of d the upper (feed) pinch controls the minimum reflux, according to the situation shown in Figure 4a. As the value of d increases, the ratio V_2^{lo}/V_0 also increases and at some point becomes equal to V_2^{up}/V_0. This point is the minimum of the relation given by Eq. 21, where both pinches are active. Further increase in d leads to the situation shown in Figure 4b, where the lower (reboiler) pinch controls, and for a given x_i the minimum vapor flow increases. The minimum value of the ratio of minimum vapor flows clearly depends on x_i, the composition in the main column where the intermediate reboiler is located. The location of the reboiler in the column is eventually determined by the bubble and dew points of the mixture, as previously discussed. In our example we have arbitrarily chosen $x_i = 0.3$ at the minimum, that is, the mole fraction of the more volatile component in the liquid at the intermediate reboiler/condenser location is 0.3. We assume that the temperature difference in the intermediate reboiler/condenser is sufficient for heat transfer at this value of x_i. The ratio of vapor flows at this point is less than unity, which means that the minimum vapor flow in the prefractionator system is lower than in a single column. However, the value of x_i cannot be treated as fixed. As the variable d increases, the pressure in the prefractionator column must decrease to keep the temperature in the reboiler constant. The intermediate reboiler/condenser temperature must decrease as well, which causes an increase in the value x_i. This is illustrated in Figure 5 by the dotted line that intersects lines of V_2^{lo}/V_0 at (arbitrarily chosen) increasing values of x_i as d increases. The dotted line intersects the V_2^{up}/V_0 line at the minimum value of the minimum vapor-flow ratio, which is less than 1. Therefore, there is a potential for energy savings when the prefractionator system is used instead of a single column.

Detailed Simulation and Discussion

Using simplified models we have shown that the prefractionator distillation system is potentially attractive when compared with a conventional distillation column. To confirm these results, we performed detailed simulations of the considered distillation systems. We selected the mixture of benzene and toluene as an example. The simulation results for the single column are shown in Figure 6. The column contains 80 theoretical stages, and the feed is introduced to stage 40. Corresponding results for the prefractionator system are shown in Figure 7. The first column in the prefractionator system contains 30 stages, with the feed stage chosen in the middle (stage number 15). The second column has 80 theoretical stages, with the feed stage fixed at number 40 and the intermediate reboiler stage varied in the simulation to achieve the desired temperature difference of at least 3.5 K. Both systems intentionally incorporate a very high number of stages, so that columns would operate close to minimum reflux. The use of excess stages allows us to keep the total number of stages and the locations of feed stages fixed as d changes. Under these conditions the reboiler duty in the single column is 4.446 kW and the sum of the duties in both reboilers in the prefractionator system shown in Figure 7 is 3.181 kW. In this example, separation in the prefractionator system requires 29% less energy than for the conventional single column. The temperatures of the heating and cooling media are the same for both cases.

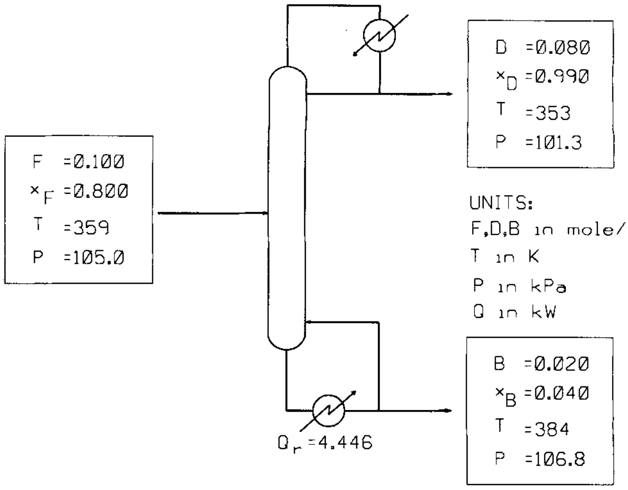


Figure 6. Simulation results for the benzene–toluene separation ($x_F = 0.8$) in a single column.

The energy savings in the prefractionator system depend on the distillate recovery from the first column (d) as previously discussed in the simplified analysis and shown in Figure 5. We examined this dependence in detail by performing a series of simulations for various d. The results are shown in Figure 8, which displays the ratio of total heat duty in the prefractionator system to the reboiler duty in the single column as a function of d. The detailed simulations yield a minimum analogous to that anticipated in Figure 5. As the recovery in the first column increases, the bottom product becomes richer in the less volatile component (toluene) and the pressure in the prefractionating column has to be decreased to keep the temperatures in both reboilers the same. As a consequence the temperature at the top of the prefractionating column decreases, which means that the intermediate reboiler needs to be moved further up in the second column to ensure the appropriate direction for heat transfer. At some

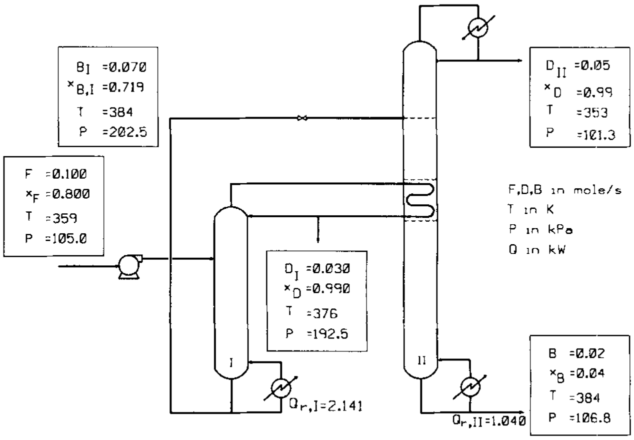


Figure 7. Simulation results for the benzene–toluene separation ($x_F = 0.8$) in a prefractionator system.

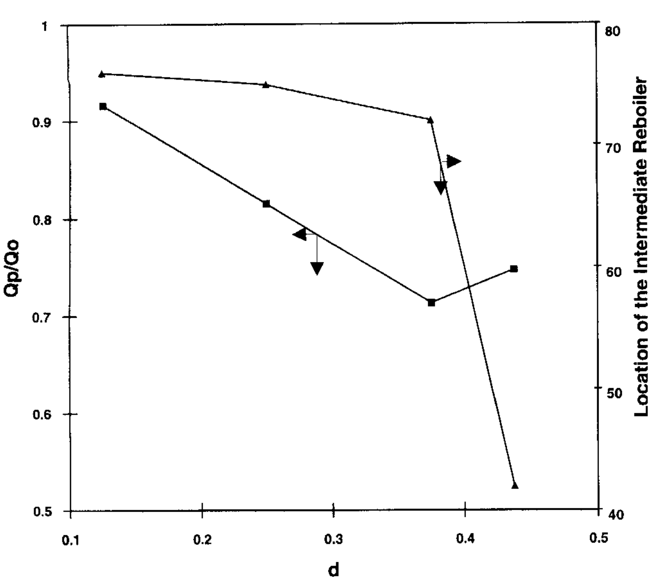


Figure 8. Ratio of the total heat duty in the prefractionator system to the single-column duty and intermediate reboiler location in the main column of the prefractionator system as a function of distillate recovery from the prefractionator.

Mixture of benzene and toluene, $x_F = 0.8$.

value of d the temperature at the top of the prefractionator becomes equal to the temperature on the feed stage in the second column. At and beyond this point, the prefractionator concept can no longer be used, since moving the intermediate reboiler farther up in the main column (and placing it in the rectifying section above the feed) will not improve the distillation performance of this column. Thus, the recovery in the first column is limited by the second law; in our example this limit occurs at $d = 0.44$. Intermediate reboiler location as a function of recovery is shown in Figure 8. In our simulations, the intermediate reboiler was located as low as possible in the main column. The minimum temperature difference across the intermediate reboiler was 3.5 K. Because of the discrete nature of the problem, this temperature difference varies from case to case between 3.5 K and 6 K. In the optimal case a 29% energy savings was achieved in comparison with a single-column separation.

This promising result gives rise to several new questions. The most important seems to be: Will the prefractionator achieve similar savings with other feed compositions, or for other mixtures? After performing some additional calculations we have concluded that the energy savings depend strongly on the composition of the feed. This conclusion can be easily explained by interpreting the process on a y-x diagram, as discussed further below. Results of simulations show that, for example, for the same mixture of benzene–toluene, but with a different feed composition, $x_F = 0.5$, only 3% energy savings can be achieved. Furthermore, the prefractionator system shown in Figure 3 is not feasible for mixtures with lower benzene content because of a decreasing temperature difference in the intermediate reboiler. As the amount of benzene in the feed decreases, the intermediate reboiler has to be located higher in the stripping section, and at some point the temperature of the reboiler/condenser becomes equal to the temperature on the feed stage of the main column. Heat transfer from the prefractionator's condenser to the stripping section of the main column is no longer possible for this feed composition, and another column configuration must be used, as explained below.

It is instructive to compare the proposed configuration (Figure 3) with a single distillation column with an additional reboiler at an intermediate location. When an intermediate reboiler is properly used in a binary distillation column, the total heat duty is not decreased, but a lower quality of heat (heat at a lower temperature) can be used in the intermediate reboiler resulting in an overall exergy savings. When all the heat is available only at one temperature level, however, the proposed prefractionator column degrades a portion of that heat to a lower level, which is still acceptable for the intermediate reboiler, while distilling at least a portion of the feed (stream F_{I} in Figure 9a) to produce a portion of the distillate product (stream D_{I}). This not only decreases the amount of the original feed to the main distillation column (stream F_{II}), but also creates a new feed (B_{I}) that is richer in the heavier component. The column with two feeds instead of one can be designed to operate more efficiently, because the irreversibilities of the distillation process can be reduced significantly (for example, see Pratt, 1967). This, coupled with the reduced quantity of feed to this column, gives an overall reduction in the energy requirements. The split of feed between streams F_{I} and F_{II}, the distillate recovery from the prefractionator column D_{I}, and intermediate reboiler location should all be optimized to achieve the highest possible energy savings. If the capital cost is included in the objective function, then the numbers of stages in all the column sections should also be varied in the optimization.

The energy savings are easily visualized on the corresponding y-x diagram shown in Figure 10a. For simplicity we assume that $F_{\mathrm{II}} = 0$, as in our simulation example. On this diagram, a single column (Figure 1a) and the main column of the prefractionataor system (Figure 3), operating close to minimum reflux, are compared. The operating lines for the single column join points B, F, and D, while the operating lines for the main column of the prefractionator system join points B, G, H, B_{I}, and D. The operating lines for the main column are drawn with a thick line in Figure 10a. An elementary material balance for the stripping section of a distillation column shows that

$$\frac{L}{V} = 1 + \frac{B}{V}, \tag{22}$$

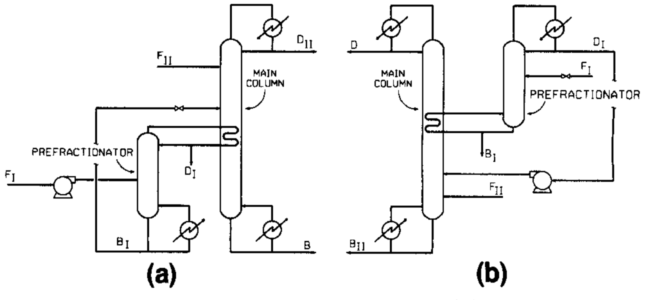
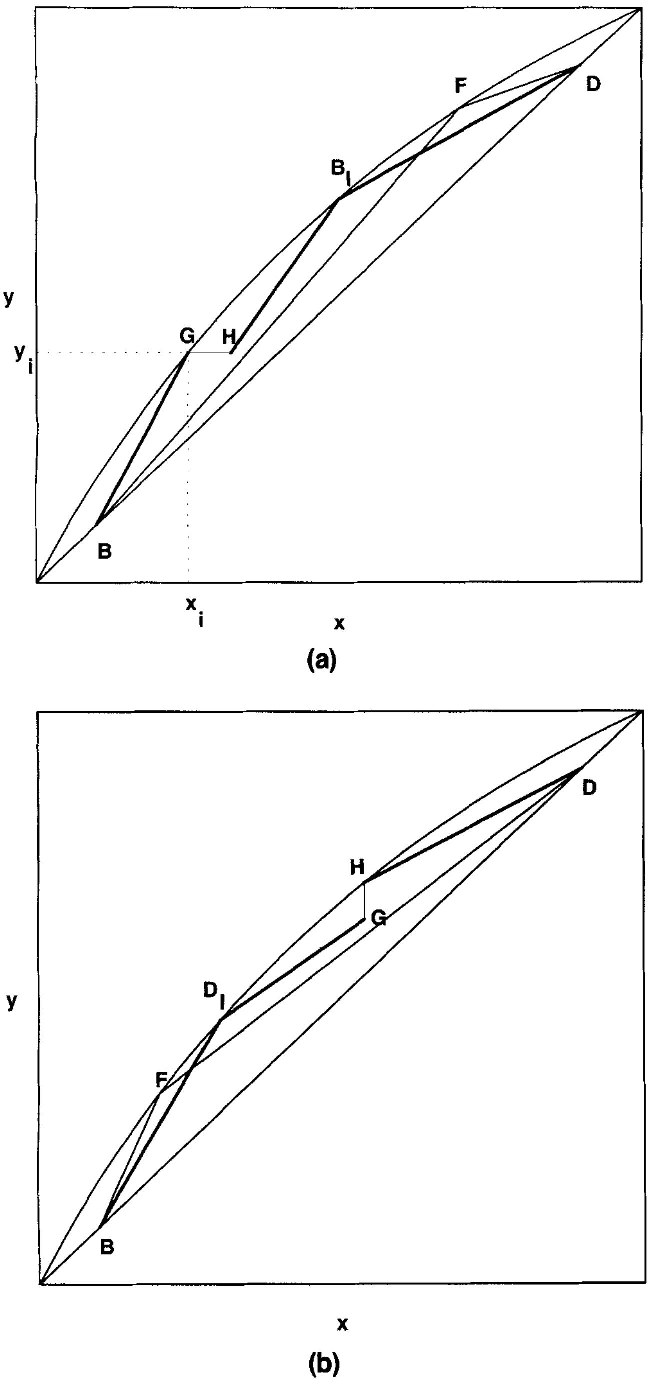


Figure 9. Prefractionator system with: (a) an intermediate reboiler; (b) an intermediate condenser.



(a)

(b)

Figure 10. Operating lines on a y-x diagram for a single column (BFD) and for the main column of the prefractionator system with: (a) intermediate reboiler ($BGHB_ID$); (b) intermediate condenser (BD_IGHD).

and the ratio L/V is the slope of line BF for the single column. Similarly, the slope of line HB_I is equal to the ratio L_2/V_2 in section 2 of the main column of the prefractionator system (section 2 is located above the intermediate reboiler, as shown in Figure 3). The previous analysis showed that the vapor flow in section 2 represents the energy requirements of the entire prefractionator system. We see that the slope of line HB_I must be greater than the slope of line BF, which means that the total vapor flow generated in the prefractionator system, and consequently its energy requirements, are smaller than those of the single column. Note that the flow rate of the bottom product is the same in both cases.

The effect of the feed composition on potential energy savings is also easy to explain on the y-x diagram. The stripping section of the single column is more inefficient for $x_F = 0.8$ than in the case of $x_F = 0.5$, since the operating line is located farther away from the equilibrium line. Therefore, there is more potential to improve the process (for $x_F = 0.8$) by utilizing the prefractionator concept (decreasing the irreversibilities in the stripping section).

An alternate prefractionator configuration to distill a feed containing a greater fraction of heavier component (toluene in our example problem) may be generated as a "mirror image" of the previous system (Figure 9b), whereby an intermediate condenser allows a portion of the condensing duty to be provided by a higher temperature cooling medium. For this new configuration all the condensing duty is available at a single low temperature, but the prefractionator converts a portion of this duty to a higher temperature while distilling a portion of the feed (stream F_I) to produce a portion of the bottom product (stream B_I). The prefractionator column must have a stripping section, to produce a heavy–rich bottom product B_I, and it has to operate at a pressure that is lower than that of the main column. The pressure of the prefractionating column must be decreased so that at least a portion of the vapor stream from the rectifying section of the main column can be condensed, providing the necessary boilup duty for the prefractionator. The condensed overhead (stream D_I) from the prefractionator is pumped to the main column. This prefractionator configuration can potentially give substantial energy savings over a single column when a feed is quite rich in the less volatile component. A construction proof is shown in Figure 10b, where (for simplicity) all the feed is assumed to enter the prefractionator column ($F_{II} = 0$). It has to be pointed out that distributing the feed stream between the prefractionator and the main column is very important. Since the feed is assumed to be a liquid, in the case where $F_{II} = 0$ the entire feed stream reduced by the first bottom product B_I has to be vaporized in the intermediate reboiler and then condensed in the top condenser of the prefractionator. This usually causes more energy than necessary to be transferred through the intermediate reboiler/condenser, resulting in the less efficient process. This situation can be prevented by distributing the feed stream between the prefractionator and the main column, which additionally improves the efficiency of the main column since a new feed to this column has been introduced.

In order to illustrate the energy-savings feature of the prefractionator scheme in Figure 9b for a feed rich in the heavy component, let us consider the separation of a mixture of benzene and toluene with a feed composition: $x_F = 0.2$. The results of simulating a single column containing 80 stages, where the feed is introduced on stage 40, are shown in Figure 11. The results for the prefractionator system with the intermediate condenser are shown in Figure 12. The prefractionator column contains 20 stages, with feed introduced at the top, and the main column has 80 stages, with intermediate condenser situated on stage 26 (stages are counted from

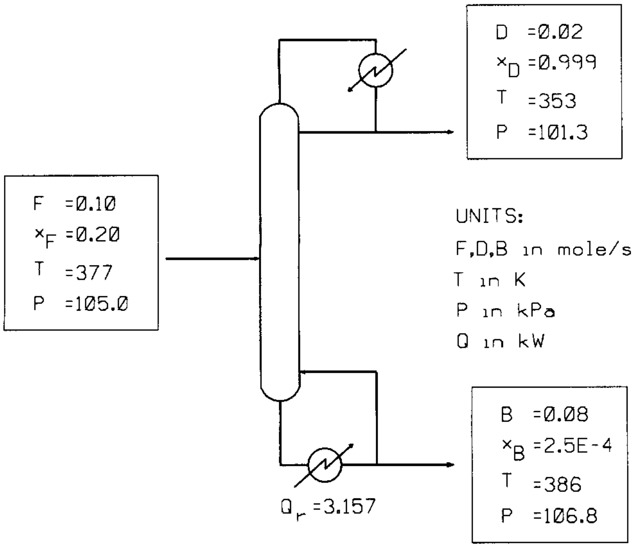


Figure 11. Simulation results for the benzene–toluene separation ($x_F = 0.2$) in a single column.

the top). A portion of the feed is introduced on stage 40, and the second feed (the distillate stream from the prefractionator column) is fed on stage 35. The prefractionator configuration with the intermediate condenser gives 9% energy savings over the conventional single column. We did not attempt to fully optimize these processes, so larger savings could probably be achieved for these prefractionator configurations. Our objective was to show that the energy required for separation can be significantly reduced by using prefractionator systems rather than a conventional single column.

A prefractionator system requires higher capital investment than a single column. The increased capital cost is needed to cover the expenses of the prefractionator column, two new heat exchangers, and the increased number of stages in the main column (where the distillation process is more reversible than in the single column). This might be paid off, however, by significant savings in the operating cost.

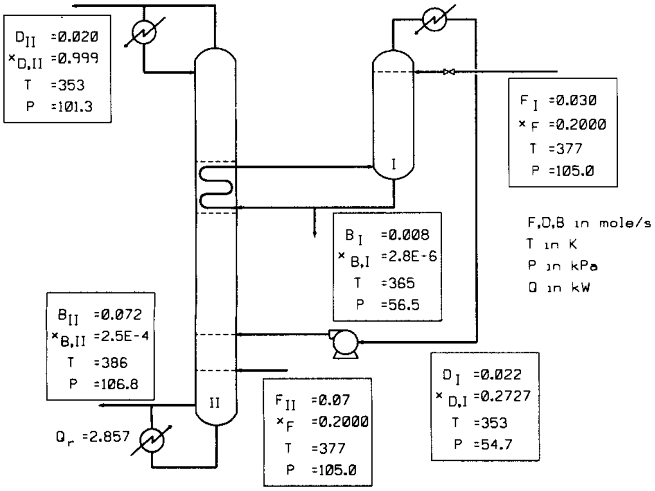


Figure 12. Simulation results for the benzene–toluene separation ($x_F = 0.2$) in the prefractionator system with intermediate condenser.

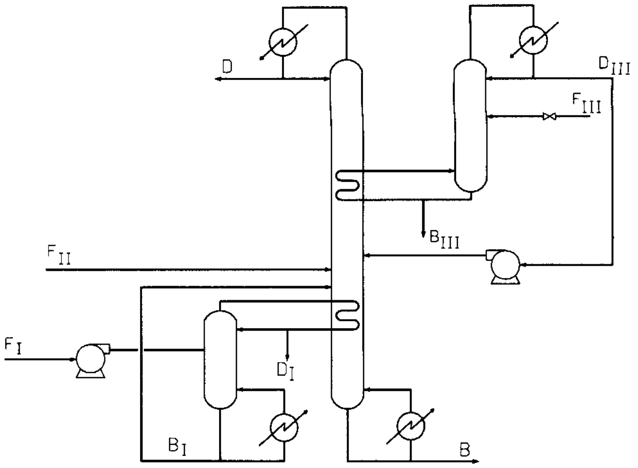


Figure 13. Prefractionator system with intermediate reboiler and intermediate condenser.

Other prefractionator configurations could meet the same objectives. One of these can be obtained by combining the discussed schemes, as shown in Figure 13. In another variation of this concept, at least a portion of a product from each prefractionating column is introduced to the other prefractionating column, rather than to the main column.

Conclusions

The total energy consumption of binary distillation is reduced when a heat-integrated, multieffect distillation system is used rather than a single column. The disadvantage of the heat-integrated system is that the temperature range between the heating medium and cooling medium increases, which can be more expensive. On the other hand, when an intermediate reboiler (or intermediate condenser) is properly applied, the total heat duty of the process remains the same as for a single column, but a part of the heating (cooling) duty can be supplied at a lower (higher) temperature than in the original single-column case. By combining these two ideas, we have obtained heat-integrated distillation systems where energy requirements are considerably reduced, but the temperatures of the utilities remain the same as for the single-column case.

Three examples of such energy-efficient distillation configurations were described. There was no trade-off between the first law heat and the second law temperature difference for these systems. Each configuration consisted of a prefractionator column heat integrated with the main column by means of an intermediate reboiler or intermediate condenser. The system with the intermediate reboiler proved to be more suitable for mixtures containing a large amount of the more-volatile component, and the system with the intermediate condenser was shown to be useful for mixtures rich in the less-volatile component. Our simulations show that an 80% benzene/20% toluene mixture can be separated with 29% less energy using the first prefractionator configuration rather than the single column. This result was achieved without splitting the feed stream between the prefractionator column and the main column, which means that there is a potential to achieve an even better result when the feed split is varied. A 9% energy savings can likewise be achieved for a 20% benzene/80% toluene mixture using the second prefractionator configuration. Obviously, the capital cost of a prefractionator configuration is higher than the cost of a single distillation column, but the total annual cost may be smaller because of quite significant energy savings.

When designing prefractionator configurations, the split of the feed stream between both columns, the product recovery from the first column (prefractionator), the number of stages in all the column sections, and the intermediate heat exchanger location that minimize the total cost of the plant should be determined.

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Notation

P = pressure
α = relative volatility
Δ = difference

Subscripts

c = condenser
F = feed
r = reboiler

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