Thermal Integration of a Distillation Column Through Side-Exchangers

Santanu Bandyopadhyay

Energy Systems Engineering and Department of Mechanical Engineering,
Indian Institute of Technology, Bombay, Powai, Mumbai 400 076, India.
Phone: +91-22-25767894, Fax: +91-22-25726875, E-mail: santanu@me.iitb.ac.in

Abstract

Invariant rectifying and stripping (IRS) curves are used for thermodynamic analysis of a distillation column. Thermal integration of a distillation column with the background process through side exchangers (side reboilers / condensers) is important for energy-efficient design of a distillation column. Thermal integration of a distillation column through side exchangers also improves the exergetic efficiency of the column. Different configurations of side-exchanger integration are evaluated based on the exergetic efficiency of the distillation process and they are also compared with feed preheating options. It has been observed that thermal integration by feed splitting is the best thermodynamic option.

1. Introduction

Distillation is one of the common unit operations used in chemical industries. Operating cost of a chemical process plant mainly depends on the operating cost associated with the separation processes. Total energy requirement of chemical process industry in United States is over 5 quad/y and distillation process alone consumes about 2.4 quad/y (Ognisty, 1995). Though distillation is energy intensive, it is not energy efficient. Therefore, it is necessary to identify different energy conservation opportunities through thermodynamic analysis of a distillation process.

Thermodynamic analysis of a distillation column is important for synthesizing and developing energy-efficient distillation processes. Thermodynamic analysis of a distillation column can be addressed through the temperature-enthalpy (T-H) curve or exergy analysis. Thermodynamic analysis of a distillation column aims at possible reduction in exergy loss, or equivalently, reduction in entropy generation. There are three major sources of entropy generation in distillation columns—heat transfer with finite temperature driving force, mixing of non-equilibrium vapour and liquid, and pressure drop across the column. Additionally, there may be entropy generation due to heat loss to the ambient from the column surface.

Bandyopadhyay et al. (1999) introduced a novel pair of temperature-enthalpy (T–H) curves, known as the Invariant Rectifying-Stripping (IRS) curves for a distillation column. The IRS curves are invariant to the column configuration (i.e., feed location in the column and number of stages) and depend only on sharpness of separation as well as operating pressure of the column. They are useful for setting quantitative targets such as minimum energy requirement, appropriate feed location, proper feed preheating, scope for side-condensers/reboilers, as well as thermo-economic optimisation of a distillation column. Exergy analysis of a distillation column often provides useful understanding and insight for energy efficient design of distillation processes. Bandyopadhyay (2002) established that the exergy loss in a distillation column may be measured directly by representing the IRS curves on Carnot factor-enthalpy (ε vs. H) diagram.

Different schemes such as proper feed location, feed preheating, side-exchangers, etc. are suggested to reduce the energy consumption in distillation. By exchanging heat with the bottom product or with any other available low-grade heat sources, thermal condition of the feed may be altered to reduce the reboiler duty. Thermal integration of a distillation column with the background process through side exchangers (side reboilers / condensers) is important for energy-efficient design of a distillation column. In refinery distillation columns, it is typically achieved through pump-around reflux. Thermal integration of a distillation column through side exchangers also improves the exergetic efficiency of the column. There are different ways an equivalent side exchanger may be integrated with the distillation column. This paper demonstrates different ways to integrate a distillation column thermally with its background process and a comparison with feed preheating is also reported. The described methodology is applicable to ideal as well as non-ideal mixtures.

2. Invariant Rectifying-Stripping (IRS) Curves

Generation procedure and physical significance of the IRS curves are briefly discussed below. Detailed derivations of these equations are given in Bandyopadhyay et al. (2004).

2.1. Generation of IRS curves

A complex column with n feeds may be decomposed into n simple columns keeping the purity and enthalpy of the end products of the i-th column same as that of the original multiple-feed column (Figure 1). For the i-th decomposed column, the distillate flow-rate (D_i), bottom flow-rate (B_i), and enthalpy difference

(Δ_i) may be determined through the overall mass, component, and energy balances ($\Delta_i \equiv Q_{ri} - Q_{ci} = D_i H_D + B_i H_B - F_i H_{Fi}$). Let H_{Ri} be the minimum condensing load (enthalpy surplus) required to cause separation from x to x_D in i-th column. Plot of enthalpy surplus as a function of the equilibrium temperature is termed the invariant rectifying (IR) curve. The overall mass, component, and energy balances for the rectifying section of the i-th column may be combined to obtain the following expression for H_{Ri}.

$$H_{Ri} = D_i\left[H_V(x_D - x)/(y - x) - H_L(x_D - y)/(y - x) - H_D\right] \tag{1}$$

Similarly, consider the stripping section of a decomposed distillation column with H_{Si} denoting the minimum reboiling load required to cause separation from x to x_B. This enthalpy deficit is then plotted as a function of the equilibrium temperature to give the invariant stripping (IS) curve.

$$H_{Si} = B_i\left[H_V(x - x_B)/(y - x) - H_L(y - x_B)/(y - x) + H_B\right] \tag{2}$$

The invariance of the IRS curves for binary systems may be proved using Gibb's phase rule. The invariant property of the IRS curves does not hold rigorously for multicomponent systems, where the concept of light and heavy keys may be employed to generate near-invariance IRS curves (Bandyopadhyay, 1999).

2.2. Order and location of feeds

Combining component, mass and energy balances around the feed stage, the following relation between the IR curve and the IS curve can be obtained.

$$H_{STi} = H_{RTi} \quad \text{at every feed stage} \tag{3}$$

where H_{RTi} (= $H_{Ri} + \Delta_i/2 + |\Delta_i/2|$) and H_{STi} (=$H_{Si} - \Delta_i/2 + |\Delta_i/2|$) are the enthalpy coordinates for the translated IR curve and translated IS curve, respectively.

Physically, equation (3) signifies the intersection of the q-line for a feed with the equilibrium curve on the x-y diagram. The intersection of the i-th q-line with the equilibrium curve is independent of the intersections for the remaining feeds. The appropriate location for i-th feed may be determined in terms of temperature (T_{Fi}) by finding the intersection of the respective translated IRS curves. The appropriate order for the feeds may be determined by simply arranging the target temperatures (e.g., feed F_2 is at a higher temperature than feed F_1 in Figure 2a).

2.3. Composite IRS curves and energy targets

The portion of the translated IR curve below T_{Fi} and the portion of the translated IS curve above T_{Fi} may be defined as the active portions of the translated IRS curves for the i-th decomposed column and consequently the i-th feed. Composite IRS curves are generated for the complex column by simply adding the enthalpy coordinates of the active translated IRS curves corresponding to all the feeds. The composite IRS curves are shown in Figure 2 for a typical two-feed column based on the translated IRS curves of the two decomposed columns. The composite IRS curves consist of $H_{RT1} + H_{RT2}$, $H_{ST1} + H_{RT2}$, and $H_{ST1} + H_{ST2}$ (Figure 2a). Thus, for section i of an n-feed column,

$$H_A = \sum_{j=1}^{n} H_{Aj} = \sum_{j=1}^{i} H_{STj} + \sum_{j=i+1}^{n} H_{RTj} \quad i = 0, 1, \dots, n \tag{4}$$

The active portions of the composite IRS curves (H_A) may be circumscribed by a right-angled trapezium. The widths of the parallel sides of the trapezium at the top and bottom define the minimum energy targets for the reboiler and condenser, respectively (see Figure 2a). Figure 2a illustrates the case where one of the intersection points of the composite IRS curves determines the pinch.

2.4. Exergy loss target

Bandyopadhyay (2003) has shown that the total exergy loss in an adiabatic distillation column may be measured by the area between the active portions of the IRS curves and the circumscribed right-angled trapezium on Carnot factor-enthalpy (ε vs. H) diagram (Figure 2b).

$$\delta Ex_T = \delta Ex_R + \delta Ex_S = \varepsilon_S H_S - \varepsilon_R H_R + \int \varepsilon \, dH_R - \int \varepsilon \, dH_S \tag{5}$$

Based on the minimum work (W_{min}) for separation and the exergy loss (δEx_T) in the column, thermodynamic efficiency of the column may be defined as follows.

$$\eta = W_{min} / (W_{min} + \delta Ex_T) \tag{6}$$

3. Side-exchanger targets

Side exchangers provide increased opportunities for heat integration and reduction in utility costs. As the IRS curves are fundamentally based on the minimum thermodynamic condition, they define the maximum heat load that can be placed on side exchangers at specified temperature levels. Figure 3a shows corners truncated out of the trapezium. The upper corner depicts the maximum scope to supply a portion ($Q_{sr, max}$) of the required heat through a side reboiler at a temperature (T_{sr}) below that of the main reboiler. Similarly a

lower corner (not shown in Figure 3a) shows the maximum potential to remove a portion ($Q_{sc,\ max}$) of the excess heat through a side condenser at a temperature (T_{sc}) above that of the main condenser. The significance of the pinch, in the context of side exchangers for distillation, may be stated as follows: no (side-)reboiling below the pinch temperature and no (side-)condensing above the pinch temperature. This is consistent with the observations of Naka et al. (1980) and Agrawal and Fidkowski (1996).

In actual practice, it is difficult to provide heat in equilibrium to a theoretical stage. In practice, it is easier to draw internal liquid or vapour stream from the distillation column, heat it with external heating medium and place it back to the distillation column. Different possible arrangements are discussed below with their exergetic efficiencies through an ideal binary example.

3.1. Base case

As an illustrative example, some simplifying assumptions are made that are consistent with the original problem description. The assumptions are no pressure drop losses, ideal vapour phase, and ideal liquid solution. Further, it is assumed that the latent heats of vaporization (λ_i) are equal for both the components and independent of temperature (over the operating temperature range of the distillation column), and the vapour pressure of a component is given by the Clausius-Clapeyron equation. These assumptions imply that the mixture to be separated is an ideal binary mixture of constant relative volatility with constant molar overflow. In this example, T_A = 353 K and T_B = 373 K where T_A and T_B are the boiling points of the low-boiling component A and the high-boiling component B, respectively. The latent heat of vaporization λ is calculated using the Clausius-Clapeyron equation from $\lambda = T_A\ T_B\ R\ \ln\ \alpha\ /\ (T_B - T_A)$ knowing the relative volatility α between components A and B. Assuming α = 2.05, λ is found to be 39.291 kJ/mol. However, it should be noted that the methodology described here is not restricted to cases with these assumptions. The described methodology is applicable to ideal as well as non-ideal mixtures. A feed of 50 kmol/h, containing 60% lighter component (A), is fed to the distillation column as saturated liquid. 98% of the lighter component is recovered at the top of the column whereas 97% of the heavier component is recovered at the bottom.

From the overall mass and material balances, the product flow rates (in kmol/h) are obtained as D = 30, and B = 20. The IRS curves of the system are shown in Figure 3a. Appropriate location for the feeds may be targeted as T_F = 359.2 K (corresponds to z = 0.6) and the minimum energy targets (in terms of condenser and reboiler loads) are determined from the widths of the circumscribing rectangle (Δ = 0, in this example). The minimum condenser and reboiler duties are equal (2.897×10^6 kJ/h), which corresponds to the minimum reflux ratio equals 1.458 and minimum reboil ratio equal to 3.687. Exergetic efficiency of the base column is found to be 56.12%. If the column operates with 20% more reflux, the exergetic efficiency decreases to 46.78%. This is considerably higher than the commonly cited thermodynamic efficiency of 10%. Therefore, design of better tray internals for lower pressure drop and the energy efficient design of the reboiler as well as condenser are to be addressed to improve the overall efficiency of the distillation column. Especially, sub-cooled reflux should be avoided to improve the thermodynamic efficiency. It has already been observed that the energy efficiency of the crude distillation unit can be improved upon by avoiding sub-cooled reflux to the main fractionation column (Joshi et al., 2001). As the primary focus is on thermodynamic efficiency and also for brevity, we assume a tall column with minimum reflux.

In case a side reboiler is incorporated in the base column, thermodynamic efficiency of the column can be improved significantly. The problem of finding location and duty of a side reboiler can be mathematically translated as cutting a rectangle with maximum area from the IRS curves represented on a Carnot factor-enthalpy (ε vs. H) diagram (as explained in Figure 3a). Maximum exergetic efficiency can be achieved when a side reboiler is placed at 363.81 K with a maximum possible duty of 1.414×10^6 kJ/h (49% of the main reboiler duty). Exergetic efficiency of the system improves to 70.92% (an improvement of 26.37% over the base column).

3.2. Drawing liquid from the column and returning at the same point after heating.

A portion of the vapour or liquid may be withdrawn from the distillation column and fed back to the column at the same point after heating it up with external utility. Drawing vapour from the column is difficult in practice. Moreover, lower value of heat transfer coefficient of the vapour stream will call for larger heat exchanger. It may also be noted that introduction of superheated vapour will have its own operational problems such as pressure drop, high velocity, product degradation due to liquid carry over, etc. For all practical purpose, it is better to withdraw liquid from the column. This is also the practice as in the cases of pump-around or pump-back reflux (Joshi et al., 2001).

For comparison purpose, the heater duty kept constant at 1.414×10^6 kJ/h. As the liquid is withdrawn at 363.81 K (corresponds to z = 0.359) and put back at the same location, the exergetic efficiency of the column improves to 70.7%. One interesting observation made during the study that the efficiency of the column is independent of the withdrawal flow rate. This implies that the flow rate of the withdrawal may be varied to adjust the appropriate approach temperature of the external heat exchanger. This is equivalent to having side reboiler at the withdrawal stage. Figure 3b shows the active portions of the IRS curves

demonstrating this. Small difference in efficiency is due to numerical integration procedure applied to calculate total exergy loss in the column. It may be noted that the withdrawal rate should be such that the resultant feed, after heating, must be saturated or two phase. Otherwise the mass transfer efficiency of the actual stages will decrease significantly as it will be doing mainly heat transfer activities.

3.3. Drawing liquid from the column and returning at different location.

A portion of the liquid withdrawn from the column may be fed back to the column at another point based on the intersection of the q-line of the heated liquid with the equilibrium curve. This is a new configuration and not analyzed before. Withdrawal of 10 kmol/h from the column at 363.81 K (corresponds to $z = 0.359$) and heating it with external utility (exchanger duty is kept equal to 1.414×10^6 kJ/h) changes the thermal condition to $q = -2.6$. Based on the q-line of this, it may be put back into the column at 370.47 K. Figure 3c shows the active portions of the IRS curves demonstrating this. It changes the internal traffic significantly and the resultant efficiency of the column comes out to be only 63.1%, significantly lower than the previous case. This is primarily due to retrograde phenomenon of internal separation. Superheated vapour (as q is negative) has to be separated inside the column such that the required separation may be obtained at the withdrawal stage. Furthermore, there are associated superheated vapour introduction problems, as highlighted before.

It may be noted that the efficiency of the column changes with liquid draw rate. Changing the withdrawal rate from 10 kmol/h to 36 kmol/h, the resultant feed becomes saturated vapour and the thermodynamic efficiency increases up to 67.11% (when located at 367.11 K). Figure 3d shows the active portions of the IRS curves demonstrating this. If the withdrawal rate is further increased, the efficiency approaches the limiting value of 70.7% and become equivalent to the previous case.

3.4. Comparison with feed preheating options

The same amount of heat may be utilized to preheat the feed. By preheating the feed by 1.414×10^6 kJ/h, the thermal energy of the feed can be changed and consequently the q value of the feed changes from 1 to 0.28. The location of the feed, in temperature scale, changes to 361.5 K. However, the change in feed quality does not change the reboiler duty alone, it also increase the condenser duty. The minimum condenser duty changes to 3.411×10^6 kJ/h, corresponds to the minimum reflux ratio of 1.894. The minimum reboiler duty changes to 1.997×10^6 kJ/h, corresponds to the minimum reboil ratio of 2.541. The preheat efficiency, defined as the portion of the heat given to feed that reduces the reboiler duty, is 63.65%. The thermodynamic efficiency of the column changes to 64.6%. Procedures developed by Bandyopadhyay et al. (2003) may be utilized to calculate the feed preheat efficiency (Figure 4a).

As proposed by Deshmukh et al (2005), feed preheat efficiency can be improved to 100% using the concept of feed splitting. Now, 23.314 kmol/h of feed may be left unchanged and fed to the column at 359.15 K. Remaining portion of the feed, i.e., 26.686 kmol/h can be preheated by the external utility heater and fed to the column at 363.73 K. This arrangement ensures 100% feed preheat efficiency (Figure 4b) as the condenser duty remain same and the reboiler duty reduced by the same amount. In this case, a maximum thermodynamic efficiency of 77.44% can be achieved.

4. Conclusion

Exergy analysis of a distillation column often provides useful understanding and insight for energy efficient design of distillation processes. This is particularly true for low-temperature processes, such as gas separation process. Exergy analysis is also useful for thermal integration of a distillation column with other unit operations. In addition to determining total exergy loss in a distillation column, it is desirable to look at its exergetic efficiency. The exergetic efficiency of a distillation column or any other unit operation can provide useful information about its potential for improvement. It is possible to have a large total exergy loss for an operation that still has fairly high exergetic efficiency. For example, even though there is a great incentive to improve further the performance of an efficient heat exchanger, handling a large mass flow, the potential for improvement is low and would be likely to require significant effort and investment. On the other hand, an operation with low exergetic efficiency and also a low exergy loss will not be worth the effort to further improve its performance. Instances where exergetic efficiencies are not very high and where exergy losses are significant provide good potential for improvement.

Thermodynamic efficiency of a distillation improves as it is thermally integrated with background process. There are two important ways a distillation column can be energy integrated: through side exchangers and through feed preheating. By proper integration of side reboiler, the thermodynamic efficiency of a column can be improved from 56.12% to 70.7% (26% improvement). On the other hand, by incorporating feed preheater with only a fraction of feed is preheated, exergetic efficiency of the column can be improved to 77.44% (38% improvement). Therefore, for the example problem, it is better to preheat the feed than to put a side reboiler. However, the improvement in thermodynamic efficiency depends on the system (equivalent to relative

volatility), feed composition, initial thermal condition of the feed, and sharpness of separation. Equivalent observations may be made for side condensers.

5. References

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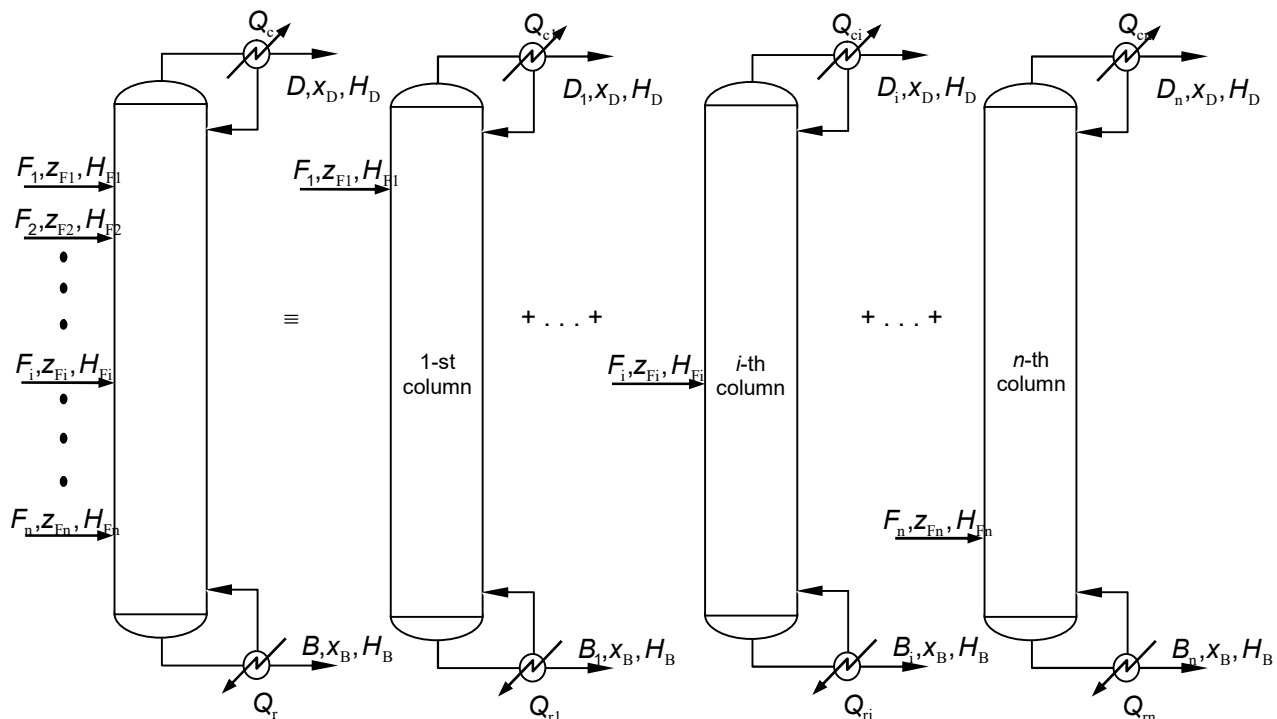
Figure 1. Decomposition of complex distillation column with n feeds into n single-feed distillation columns.
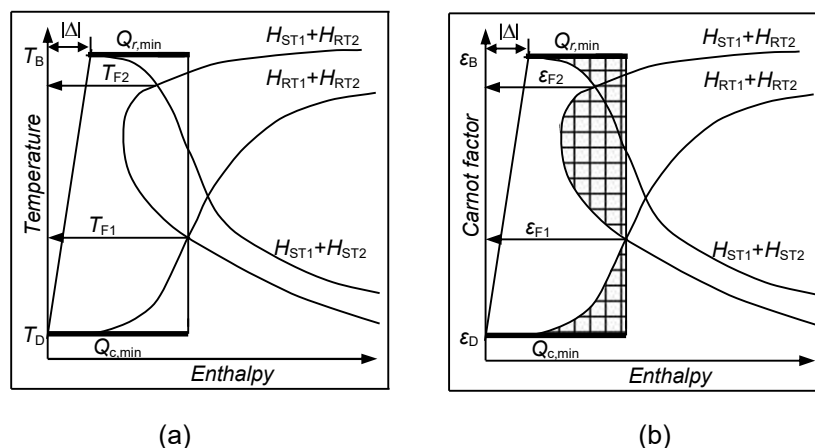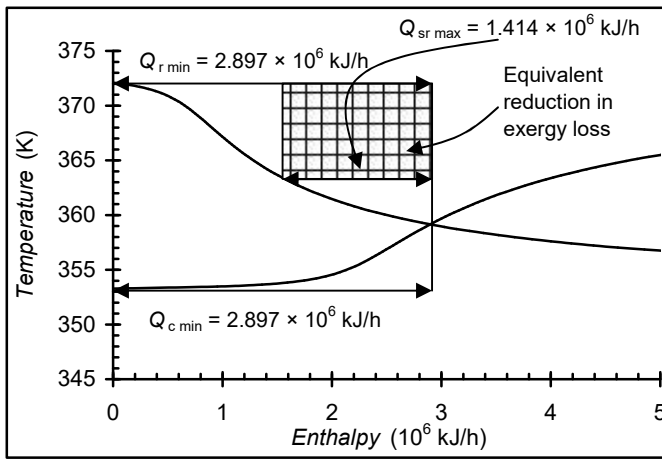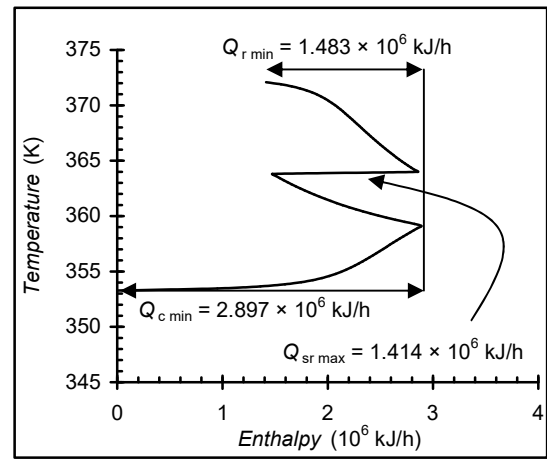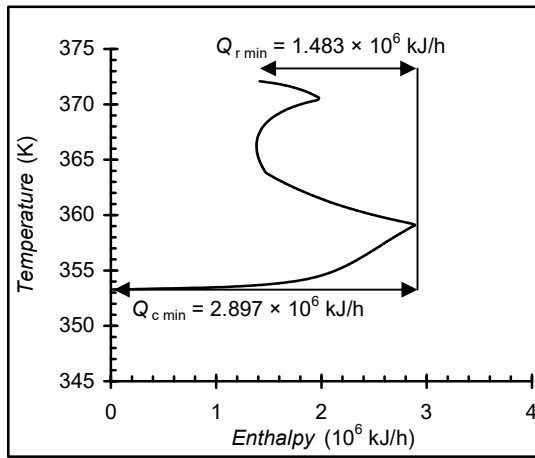

(a)   (b)

Figure 2. Composite IRS curves for a two-feed column: (a) targeting feed location and minimum energy requirement; (b) targeting exergy loss in a distillation column.
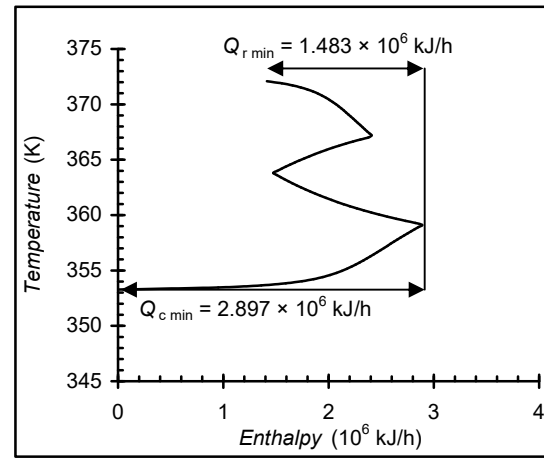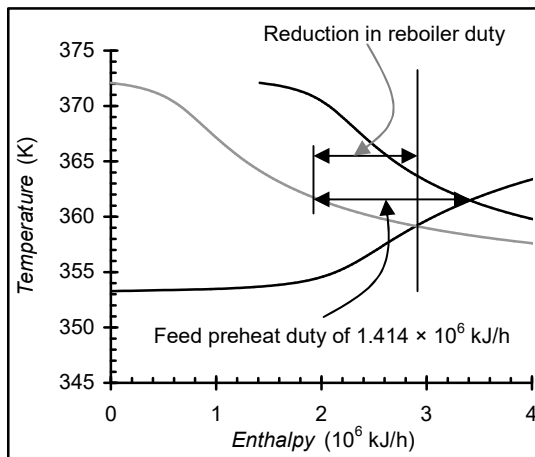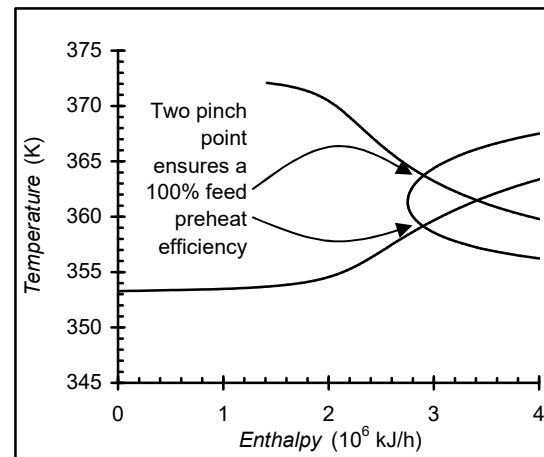
Figure 3. Active IRS curves for different way of incorporating side reboiler: (a) side reboiler targets; (b) withdrawing liquid and returning at the same location; (c) withdrawing 10 kmol/h of liquid and returning at different location; and (d) withdrawing 36 kmol/h of liquid and returning at different location.

Figure 4. Feed preheating: (a) entire feed is preheated and (b) 100% feed preheat efficiency by feed splitting.