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Effect of feed on optimal thermodynamic performance of a distillation column

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Abstract

Invariant rectifying-stripping (IRS) curves are presented for thermodynamic analysis of distillation column. The IRS curves are invariant to the column configuration (i.e., feed location and total number of stages) and therefore, they are useful to set quantitative targets such as minimum energy requirement (minimum condenser and reboiler duties), appropriate feed location, proper feed preconditioning, scope for side-condensers/reboilers, as well as thermo-economic optimization of distillation column.

In this paper, the relation between the IRS curves and exergy loss in an adiabatic distillation column is established through a simple graphical representation. A general methodology is put forward to analyze the optimal thermodynamic performance of a distillation column. Effects of feed composition, thermal condition of the feed, relative volatility of components, and sharpness of separation on the thermodynamic efficiency of a distillation column are presented in this study. The methodology, described in this paper, is equally applicable to real systems, with or without tangent pinch. A general principle for feed preheating and side-exchanging is also developed.

Keywords: Pinch analysis; Exergy analysis; IRS curves; Distillation column; Feed condition; Optimization

1. Introduction

Thermodynamic analysis of a distillation column is important for synthesizing and developing energy efficient distillation processes. It allows the thermodynamic efficiency of the process to be quantified, regions with poor energy efficiency to be identified, and the thermodynamic targets to be defined. 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 vapor and liquid, and pressure drop across the column. Furthermore, there may be entropy generation due to heat loss to the ambient from the column surface. There are two different ways of analyzing distillation column thermodynamically, viz., exergy analysis and an approach based on the temperature–enthalpy (*T*–*H*) curve.

Exergy analysis of a distillation column often provides useful understanding and insight for energy efficient design of distillation processes [1–4]. 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 a 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.

Several studies show the application of exergy analysis to improve the thermodynamic efficiency of a distillation column [1–4]. Atkinson [1] has developed a graphical representation of exergy loss in a distillation column. Ratkje et al. [2] have analytically shown that entropy generation for a distillation column is at a minimum when the driving force for separation is distributed uniformly along the column. Taprap and

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Nomenclature

B	bottom product molar flow
D	distillate molar flow
Ex	exergy
F	feed molar flow
H	enthalpy
L	liquid molar flow
Q	heat duty
S	entropy
T	temperature
V	vapor molar flow
x	mole fraction in liquid
y	mole fraction in vapor
z	mole fraction in feed

Greek letters

α	relative volatility
Δ	enthalpy difference defined in Eq. (11)
ε	Carnot factor

Subscripts and superscripts

B	bottom product
c	condenser
D	distillate
F	feed
L	liquid
min	minimum
r	reboiler
R	rectifying
RT	rectifying (translated)
S	stripping
ST	stripping (translated)
V	vapor
$*$	equilibrium condition

Ishida [3] have presented different exergy losses in a distillation column on energy-utilization diagrams. These diagrams identify the amount of energy transformation and exergy loss of individual process steps. Agrawal and Herron [4] have given equations to quantify thermodynamic efficiency of a distillation column that separates an ideal binary mixture, with constant relative volatility, into pure components.

The *T–H* curve for a distillation column quantitatively address the energy saving potential for possible stand-alone modifications as well as process integration. The *T–H* curve for a binary distillation column at minimum thermodynamic condition (MTC) is generated by solving the coupled mass and enthalpy balances for the reversible separation scheme [5–11]. The MTC for distillation is defined as reversible operation of a column without any entropy generation. It corresponds to a column with infinite stages having a side exchanger at every stage without any pressure drop and any heat loss through its surface across the column (as discussed in detail by Bandyopadhyay et al. [8]). In contrast to binary distillation, the sharpness of separation is generally limited in multicomponent reversible distillation [6] and it is impossible to device a reversible separation scheme for many practical multicomponent separations [10]. The limitations of sharpness of reversible multicomponent distillation may be overcome, for the purpose of generation of the *T–H* curve, by using the pseudo-binary concept of light and heavy key [6–9,11]. Dhole and Linnhoff [7] called the *T–H* curve of a distillation column the column grand composite curve (CGCC) and described a procedure for generating it for a simulated base-case distillation column. The generation procedure inherently accounts for the inevitable losses like feed loss, pressure loss, loss due to chosen configuration, and sharp separation loss. The shortcomings in the thermodynamic analysis based on the CGCC are discussed in detail by Bandyopadhyay et al. [9].

Bandyopadhyay et al. [9] introduced a novel pair of *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 (for condenser and reboiler), appropriate feed location, proper feed preconditioning, scope for side-condensers/reboilers, as well as thermo-economic optimization of distillation column design [9,11].

These two seemingly different approaches, i.e. the approach based on *T–H* curve and exergy analysis of a distillation column are closely related. In this paper, the relation between the IRS curves and exergy loss in a distillation column is established analytically and a simple graphical representation for exergy loss in an adiabatic distillation is developed by representing the IRS curves on Carnot factor–enthalpy (ε vs. H) diagram. Therefore, IRS curves can be used for setting quantitative energy targets as well as exergy analysis of a distillation column.

The impact of feed on optimal thermodynamic performance of a distillation column is not well studied. Typically, feed to a distillation column originates from another distillation column or other unit operations. Knowing the effect of thermal condition of the feed on the optimal thermodynamic performance of a distillation column, thermodynamic efficiency of the distillation column, as well as that of the overall process, may be improved upon by adjusting the feed at its optimal condition from the preceding unit operation. Agrawal and Herron [4] recently studied the impact of feed condition on thermodynamic efficiency of distillation column. However, their equations and methodology are limited to ideal binary mixture with constant relative volatility and are not even applicable for systems where the variation in relative volatility varies moderately from top to bottom of the column. In this paper, a general methodology is presented to analyze the thermodynamic performance of a distillation column using the IRS curves to address the dependency of thermodynamic efficiency of a distillation column on feed composition, thermal condition of the feed and relative volatility of components. Moreover, the assumption of pure component recovery made in the earlier study [4] is relaxed in this paper and the effect of sharpness of separation on thermodynamic performance of a distillation column is also studied. The methodology described in this paper is equally applicable to non-ideal systems with or without tangent pinch in either section of the column.

2. The invariant rectifying-stripping curves

The IRS curves are briefly discussed below based on a derivation for a simple distillation column (with a single feed and two products) at the MTC. It should be noted that the operating curve coincides with the equilibrium curve at MTC.

2.1. Invariant rectifying curve

For the envelope of a rectifying section of a distillation column (as shown in Fig. 1a), the overall mass balance, component balance, and energy balance may be combined to yield the following [9]:

$$H_R = \int dQ = D\left[H_V \frac{x_D - x^*}{y^* - x^*} - H_L \frac{x_D - x^*}{y^* - x^*} - H_D\right] \quad (1)$$

The quantity H_R signifies the minimum condenser load required to carry out a separation from x^* to x_D. This enthalpy surplus may be plotted as a function of temperature to give a T vs. H_R curve or it may be plotted as a function of Carnot factor ($\varepsilon = 1 - T_0/T$) to give a ε vs. H_R curve, which may be termed as the invariant rectifying curve. A typical invariant rectifying curve is shown in Fig. 1b.

Energy balance of a rectifying section may be written as

$$V_{\min} \int dH_V = L_{\min} \int dH_L + D \int dH_D + \int dQ \quad (2)$$

Differentiating Eq. (2), dividing through out by T, and integrating the resultant, we get

$$V_{\min} S_V = L_{\min} S_L + D S_D + \int \frac{dQ}{T} \quad (3)$$

Note that the basic thermodynamic relation $dH = T\,dS + V\,dP + \sum N_i\,d\mu_i$ is utilized to derive the above equation. Eq. (3) signifies the entropy balance of the rectifying section, shown in Fig. 1a and there is no entropy generation. This proves that the rectifying section is reversible and corresponds to MTC.

Denoting Ex ($= H - T_0 S$) as the availability function, the exergy balance for the rectifying section may be written as

$$\mathrm{Ex}_V = \mathrm{Ex}_L + \mathrm{Ex}_D + \int \left(1 - \frac{T_0}{T}\right) dQ \quad (4)$$

Instead of using distributed condensing load to perform the separation, usually only one condenser is used in simple distillation column (adiabatic distillation column). Use of single condenser results in entropy generation in the column. The exergy loss in the rectifying section of an adiabatic column may be obtained as

$$\begin{aligned}\delta \mathrm{Ex}_R &= \int \left(1 - \frac{T_0}{T}\right) dQ - \left(1 - \frac{T_0}{T_D}\right) \int dQ \\ &= \int \varepsilon\, dH_R - \varepsilon_D H_R \end{aligned} \quad (5)$$

Elegant geometric interpretation of Eq. (5) may be obtained on ε vs. H_R diagram as shown in Fig. 1b. The shaded area

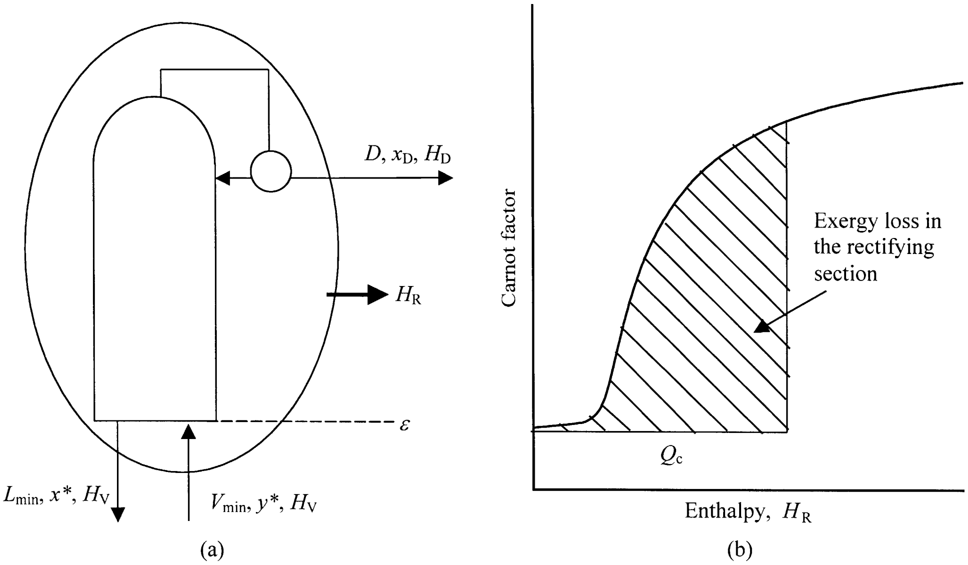


Fig. 1. Invariant rectifying curve: (a) rectifying section of a column at MTC; (b) typical invariant rectifying curve and associated exergy loss in the rectifying section of an adiabatic column.

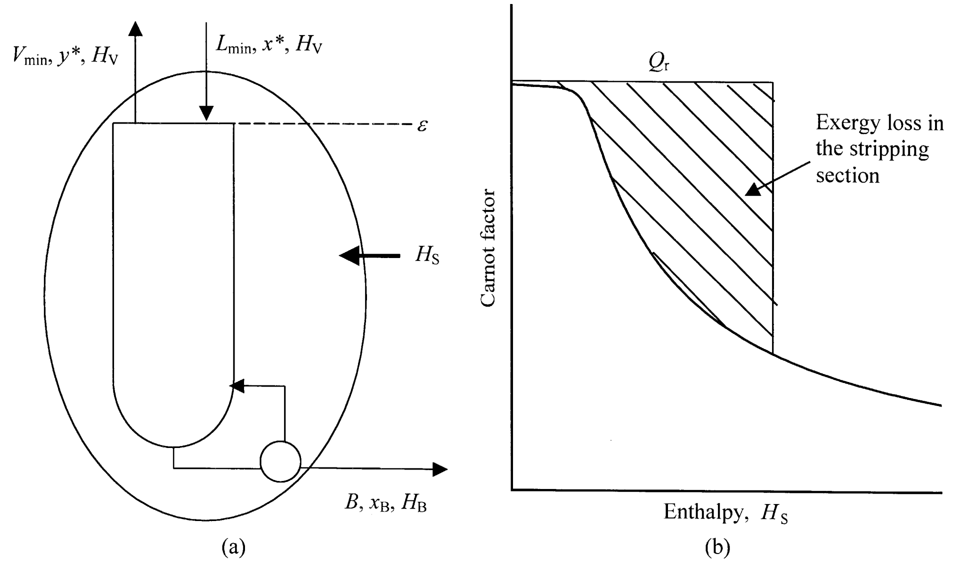


Fig. 2. Invariant stripping curve: (a) stripping section of a column at MTC; (b) typical invariant stripping curve and associated exergy loss in the stripping section of an adiabatic column.

represents the exergy loss in the rectifying section for a given condenser duty (Q_c).

2.2. Invariant stripping curve

In a similar fashion as the rectifying section, the overall mass balance, component balance, and energy balance for the envelope of a stripping section of a distillation column (as shown in Fig. 2a), may be combined to yield the following [9]:

$$H_S = \int \mathrm{d}Q = B\left[H_V \frac{x^* - x_B}{y^* - x^*} - H_L \frac{y^* - x_B}{y^* - x^*} + H_B\right] \quad (6)$$

The quantity H_S signifies the minimum reboiling load necessary to carry out a separation from x^* to x_B. This enthalpy deficit is plotted as a function of Carnot factor to yield a ε vs. H_S curve which may be termed as the invariant stripping curve. Fig. 2b shows a typical invariant stripping curve.

Following the procedure as described earlier, the entropy balance and the exergy balance of the stripping section may be obtained as

$$L_{\min} S_L + \int \frac{\mathrm{d}Q}{T} = V_{\min} S_V + B S_B \quad (7)$$

$$\mathrm{Ex}_L + \int \left(1 - \frac{T_0}{T}\right) \mathrm{d}Q = \mathrm{Ex}_V + \mathrm{Ex}_B \quad (8)$$

The above two equations prove that the stripping section also corresponds to MTC.

Instead of using distributed reboiling load to perform the separation, usually only one reboiler is used in an adiabatic distillation column. Use of single reboiler results in exergy loss in the column. The exergy loss in the stripping section of an adiabatic column may be obtained as

$$\delta \mathrm{Ex}_S = \left(1 - \frac{T_0}{T_S}\right) \int \mathrm{d}Q - \int \left(1 - \frac{T_0}{T}\right) \mathrm{d}Q = \varepsilon_S H_S - \int \varepsilon \, \mathrm{d}H_S \quad (9)$$

The shaded area (shown on ε vs. H_S curve in Fig. 2b) represents the exergy loss in the stripping section for a given reboiler duty (Q_r).

The invariant rectifying curve and the invariant stripping curve may be plotted on the same T–H or ε–H axes. In this paper, the IRS curves are drawn on same ε–H axes to quantify the exergy loss in a distillation column (see Fig. 3).

2.3. IRS curves and their physical significance

Physically, the IRS curves correspond to the enthalpy surpluses and deficits for the rectifying and stripping sections respectively, for all possible values of reflux and reboil after neglecting the effect of the feed.

A binary two-phase system has exactly two degrees of freedom as per Gibbs phase rule. On specifying the operating pressure and the separation, the system becomes deterministic. Therefore, H_R and H_S are functions of temperature only. In other words, the IRS curves are invariant to the feed location and the operating reflux for a distillation system whose operating pressure and separation are specified.

For reversible multicomponent distillation, the degrees of freedom are still 2 [12]. By arguments analogous to those for the binary case, the system becomes deterministic. The IRS curves are invariant to the feed location and the operating reflux on specifying the operating pressure and the separation.

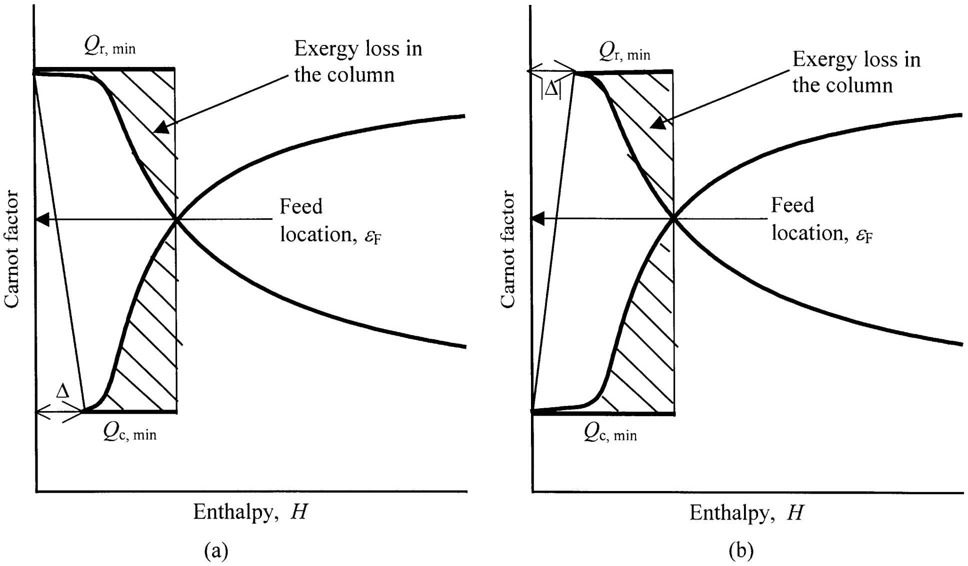


Fig. 3. Appropriate feed location (ε_F), minimum utility requirement (Q_r and Q_c) and targeting exergy loss of a column through translated IRS curves: (a) $\Delta \geq 0\,(Q_r \geq Q_c)$; (b) $\Delta < 0\,(Q_r < Q_c)$.

The sharpness of separation is generally limited in reversible multicomponent distillation [6,10]. However, this limitation can be overcome during the generation of IRS curves using the pseudo-binary concept of a light and heavy key [6–9,11] that defines a practical near-MTC [7]. The invariant property of the IRS curves does not hold rigorously for multicomponent systems because the distribution of the mole fractions of the components depends on the operating reflux of the column. However, the IRS curves for any pseudo-binary system can be taken to be invariant of the number of stages and the feed location for targeting purposes [9,11]. Essentially, IRS curves for establishing targets in multicomponent systems must be generated through a simulation with a high number of stages (i.e., at a low reflux ratio). However, for brevity, case studies in this paper are restricted to binary systems only.

2.4. Feed location criterion and translated IRS curves

The invariant rectifying curve (1) and the invariant stripping curve (6) are not independent. These curves are related through the overall material, component, and enthalpy balance of the entire distillation column [9]:

$$H_S = H_R + F\left[H_L\frac{z_F - y^*}{y^* - x^*} - H_V\frac{z_F - x^*}{y^* - x^*} + H_F\right] + \Delta \quad (10)$$

where the parameter Δ is defined in terms of the overall enthalpy balances for the whole column:

$$DH_D + BH_B - FH_F = Q_r - Q_c \equiv \Delta \quad (11)$$

Whenever the feed stage is pinched or the column is operating at the MTC, the material balance, component balance and enthalpy balance at a feed stage may be combined as follows:

$$H_L\frac{z_F - y_F^*}{y_F^* - x_F^*} - H_V\frac{z_F - x_F^*}{y_F^* - x_F^*} + H_F = 0 \quad (12)$$

On substituting Eq. (12) into Eq. (10), the following relation is obtained:

$$H_S = H_R + \Delta \quad \text{at the feed stage} \quad (13)$$

The invariant rectifying curve and/or the invariant stripping curve may be translated horizontally in accordance with Eq. (13). Depending on the sign of Δ (as defined in Eq. (11)), the translations may be conveniently classified into two cases:

(a) If $\Delta \geq 0$ (i.e., $Q_r \geq Q_c$), then the invariant rectifying curve is translated to the right by Δ with no shift in the invariant stripping curve (Fig. 3a).
(b) If $\Delta < 0$ (i.e., $Q_r < Q_c$), then the invariant stripping curve is translated to the right by $|\Delta|$ with no shift in the invariant rectifying curve (Fig. 3b).

Mathematically, the horizontal translations of the IRS curves may be represented as

$$H_{RT} = H_R + \tfrac{1}{2}\Delta + |\tfrac{1}{2}\Delta| \quad (14)$$

$$H_{ST} = H_S - \tfrac{1}{2}\Delta + |\tfrac{1}{2}\Delta| \quad (15)$$

Eqs. (13)–(15) may be combined to obtain $H_{ST} = H_{RT}$ at the feed stage. Thus, the important conclusion is that the point of intersection of the translated IRS curves (as shown in Fig. 3) defines the target temperature for locating the feed (ε_F). An algorithmic procedure to appropriately locate feed in a finite column is described by Bandyopadhyay et al. [9].

2.5. Total exergy loss

If the feed is properly located at ε_F, then the total exergy loss in the entire column may be established as follows: the portion of the invariant rectifying curve below ε_F and the portion of the invariant stripping curve above ε_F may be circumscribed by a right-angled trapezium. Then, the pinch on the IRS curves is defined as the point touching the vertical side of the trapezium. The widths of the parallel sides of the trapezium at the top and bottom define the absolute minimum energy targets for the reboiler ($Q_{r,\min}$) and condenser ($Q_{c,\min}$), respectively (see Fig. 3). Following Eqs. (5) and (9), total exergy loss in the column is given by the area between the active portions of the IRS curves and the circumscribed right-angled trapezium (see Fig. 3):

$$\begin{aligned}\delta \mathrm{Ex}_T &= \delta \mathrm{Ex}_R + \delta \mathrm{Ex}_S \\ &= \varepsilon_S H_S - \varepsilon_R H_R + \int \varepsilon \, \mathrm{d}H_R - \int \varepsilon \, \mathrm{d}H_S \qquad (16)\end{aligned}$$

Based on the minimum work ($W_{\min}$) for separation and the exergy loss ($\delta \mathrm{Ex}_T$) in the column, thermodynamic efficiency of the column may be defined as follows:

$$\eta = \frac{W_{\min}}{W_{\min} + \delta \mathrm{Ex}_T} \qquad (17)$$

For a given feed condition, exergy loss in the column and the thermodynamic efficiency can be obtained from the IRS curves as given in Eqs. (16) and (17), respectively. Effect of feed condition on the thermodynamic performances of a distillation column can be obtained from repetitive study of the IRS curves. Results are presented and discussed in the following section.

Fig. 3 illustrates the case where the intersection point of the translated IRS curves determines the pinch. This is often the case. Mathematically, it requires the IRS curves to be monotonic in nature. However, exceptions exist as for the non-ideal systems where tangent pinch (corresponds to *x*–*y* diagram) exists. For non-ideal systems tangent or near-tangent pinch exists in either section of the column and for multicomponent systems pinch zones are there in both rectifying and stripping section of the column. Consequently, the IRS curves are not monotonic.

3. Effect of different feed conditions

The effect of different feed conditions on thermodynamic efficiency of a distillation column are discussed in this section. For this purpose certain simplified assumptions are made. No pressure drop across the column and no heat loss to the ambient are assumed. Vapor and liquid phases are assumed to be ideal with constant latent heat. Clausius–Clapeyron equation are assumed to be valid for calculating vapor pressure of the system. Further, the distillation column is always taken to be pinched. However, it should be noted that for real systems, including non-ideal systems, even if there exists a tangent pinch or near-tangent pinch in either section of the column, equations derived in the earlier section are valid.

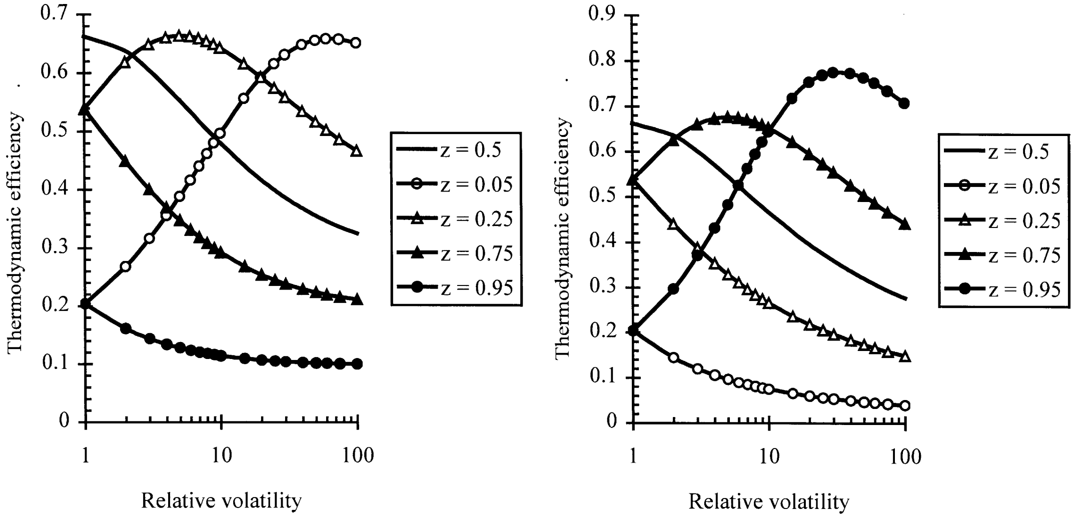
The translated IRS curves for a simplified system with a given feed condition and relative volatility are generated through Eqs. (1), (2), (14) and (15). Total exergy loss of the column is then calculated via numerically integrating the translated IRS curves as indicated by Eq. (16). Thermodynamic efficiency of the column is a consequence of Eq. (17). Variation in thermodynamic efficiencies for different feed conditions are discussed as follows.

3.1. Saturated feed

Thermodynamic efficiency of a distillation column for different values of relative volatility and saturated feed composition are shown in Fig. 4. Recovery of components are assumed to be 99% in this section.

Thermodynamic efficiency of a distillation column is commonly cited to be about 10%. In this exercise, thermodynamic efficiency of a distillation column is in the range of 26.6–67.7% for feed composition lying in between 25 and 75% and relative volatility less than 10. The major sources of thermodynamic irreversibility are due to over reflux (more than the minimum reflux for a finite column) provided in the column, pressure drop across the column and temperature driving force provided at reboiler as well as condenser. It should be noted that for 20% more reflux, exergy loss in the column may increase by only 20% and the efficiency of the column may still be higher than the reported value of 10%. For example, a pinched distillation column, processing saturated liquid feed with 50% light component at relative volatility of 2, has an efficiency of 51.8%. Thermodynamic efficiency of the same column reduced to 43.2% when 20% more reflux is used. 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.

For a very low relative volatility ($\alpha \to 1$), minimum duties for the reboiler and the condenser increase sharply, whereas the efficiency approaches a limiting value. This limiting value depends on the feed composition but is independent of the thermal condition of the feed (Fig. 4). In such circumstances, approach temperatures for the reboiler and condenser play a dominant role in defining the efficiency of the column. Similar observations were also made by Agrawal and Herron [4]. Thermodynamic efficiency of a distillation column processing saturated liquid feed with composition z is qualitatively comparable, but not exactly equal numerically, with the thermodynamic efficiency of a column processing saturated vapor feed with composition $1 - z$.



(a) (b)

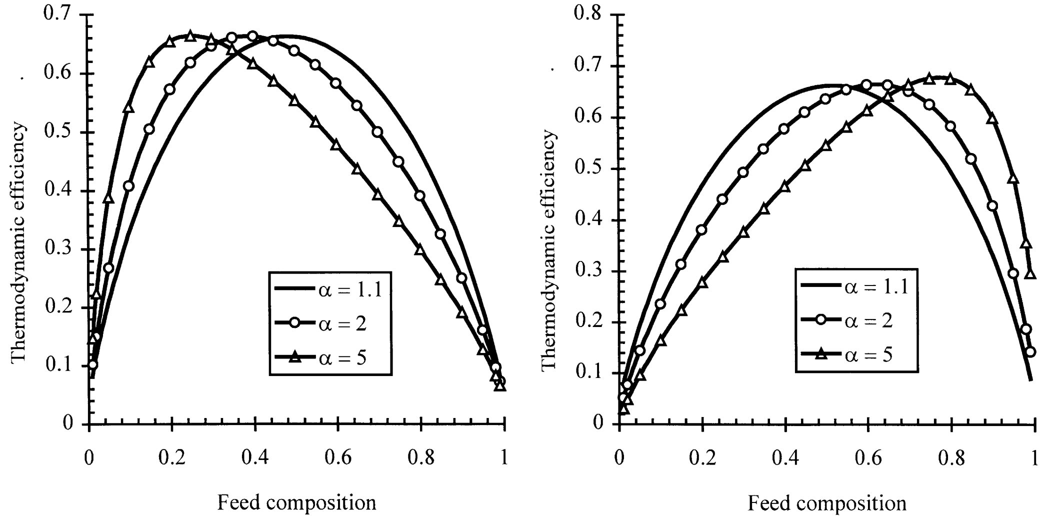
Fig. 4. Thermodynamic efficiency of a distillation column, processing saturated feed for different relative volatility: (a) saturated liquid feed; (b) saturated vapor feed.

This difference is primarily due to difference in recovery (99% recovery is assumed in this paper compared to 100% recovery as assumed by Agrawal and Herron [4]).

For saturated liquid feed with more light component ($z \geq 0.5$) and saturated vapor feed with more heavy component ($z \leq 0.5$), the efficiency reduces monotonically with relative volatility (Fig. 4). Whereas for saturated liquid feed with more heavy component ($z < 0.5$) and for saturated vapor feed with more light component ($z > 0.5$), the efficiency goes through a maximum. As concentration of the lighter component decreases in the saturated liquid feed or concentration of the heavier component decreases in the saturated vapor feed, the optimum shifts towards higher relative volatility.

Variation in thermodynamic efficiency for different feed compositions are shown in Fig. 5. For saturated liquid feed with more heavy component or saturated vapor feed with more light component, increase in relative volatility improves the thermodynamic efficiency of the column.

These observations lead to the fact that for a given composition and thermal condition, the relative volatility may be changed by changing the operating pressure to increase the efficiency of distillation. Change in operating pressure is specially important for sub-ambient distillation processes.



(a) (b)

Fig. 5. Thermodynamic efficiency of a distillation column, processing saturated feed for different feed compositions: (a) saturated liquid feed; (b) saturated vapor feed.

For saturated liquid feed with more light component (or saturated vapor feed with more heavy component), increase in operating pressure may be encouraged for cryogenic distillation. Higher operating pressure results in lower entropy generation due to lower pressure drop and reduction in refrigeration load. It should be noted that it may be possible to manipulate relative volatility to a small extent by changing the operating pressure of a distillation column, large changes in relative volatility are normally difficult to achieve. It may be worth emphasizing that in most gas separation processes the variations of distillation efficiency with operating pressure is of utmost importance. Depending on the thermal condition of the feed, the operating pressure of distillation may be adjusted. However, cost–benefit analysis should be performed before implementing such modifications.

The optimal relative volatility for different compositions of the saturated feed are shown in Fig. 6. It may be anticipated that for a given composition and relative volatility, the thermal condition of the feed may be changed to optimize the thermodynamic efficiency. This issue is discussed next.

3.2. Two-phase feed

Thermal condition of the feed may be varied to optimize the thermodynamic efficiency of a distillation column. Sub-cooled liquid or super-heated vapor cannot be fed reversibly in a distillation column [11]. Therefore, in this study, the thermal condition of the feed is restricted within these two saturated limits ($1 \geq q \geq 0$).

Variation of thermodynamic efficiency with relative volatility for a given feed condition ($z = 0.25$) is shown in Fig. 7. Distillation column processing saturated liquid feed (for $z = 0.25$) is thermodynamically more efficient than a column processing saturated vapor feed. Thermodynamic efficiency goes through a maximum for saturated liquid feed. It may be noted that the thermodynamic efficiency for a two-phase feed is monotonically increasing (Fig. 7) and the thermodynamic efficiency of a two-phase feed is always higher than that of the saturated feed beyond a certain relative volatility (for $\alpha > 5$ in this example with $z = 0.25$). In other words, optimal thermal condition of the feed, below a certain relative volatility, corresponds to a condition such that it is either saturated liquid or saturated vapor depending on the feed composition. Below a certain relative volatility, saturated liquid feed is more efficient when it contains less lighter component. On the other hand, saturated vapor corresponds to the optimal thermal condition for feed with more lighter component.

Variation of optimal thermal conditions of the feed for different feed compositions and relative volatility are shown in Fig. 8. Feed with more light component should be vaporized more to improve thermodynamic efficiency of the column. Optimal thermal condition for the feed with predominantly

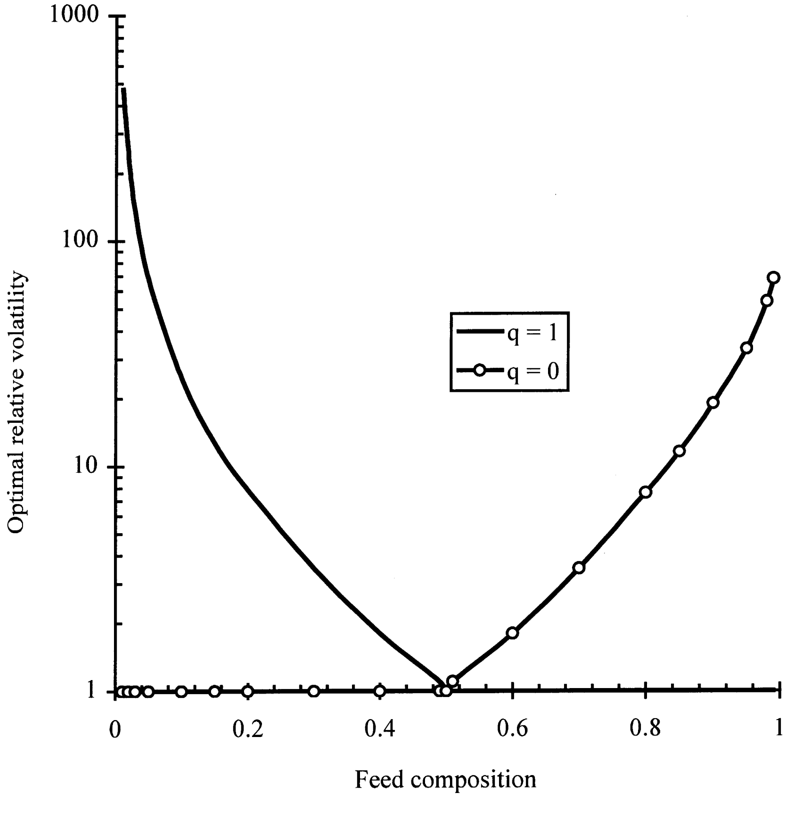


Fig. 6. Optimal relative volatility for different saturated feed compositions.

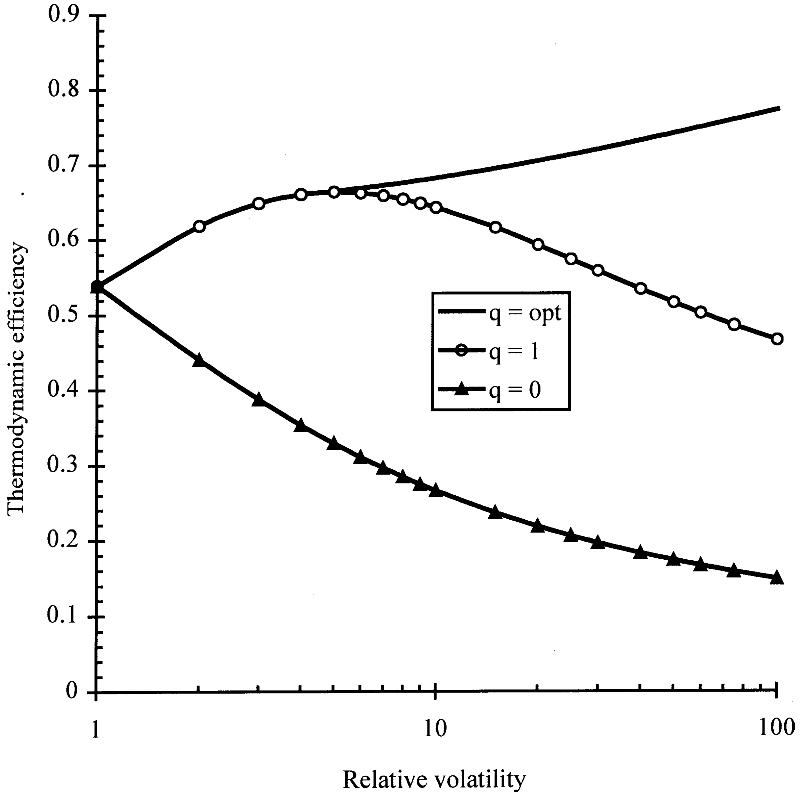
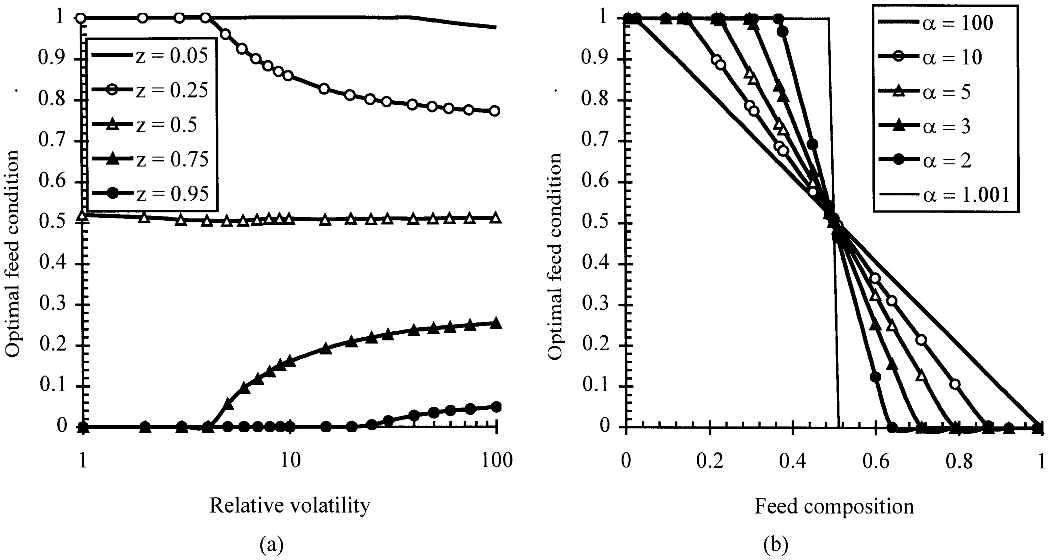


Fig. 7. Variation of thermodynamic efficiency with relative volatility for different thermal conditions of feed ($z = 0.25$).

lighter component decreases with increasing relative volatility. On the other hand, optimal thermal condition for the feed with predominantly heavier component increases with increasing relative volatility. For a given relative volatility, feed composition and optimal thermal condition of the feed closely bear a linear relationship (Fig. 8b). In many of the distillation columns, such as naphtha stabilizer column, feed-bottom exchangers are provided to increase the energy efficiency of the process. Based on the relative volatility and the feed composition, thermal condition of the feed may easily be manipulated by the designer to achieve higher thermodynamic efficiency of the systems.

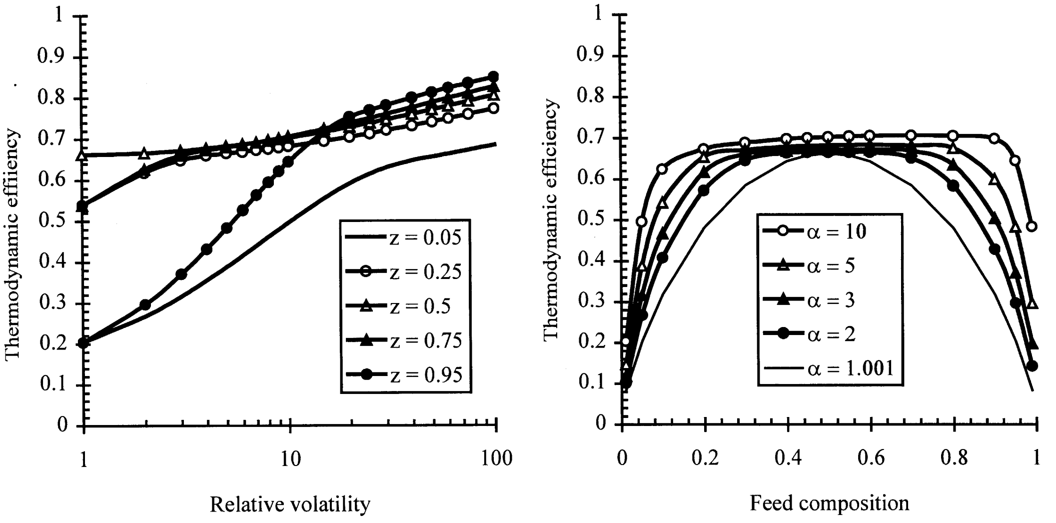
Fig. 9 shows the variation of thermodynamic efficiency with relative volatility and feed composition for optimal



(a)

(b)

Fig. 8. Optimal thermal condition of the feed for different: (a) relative volatilities; (b) feed compositions.



(a)

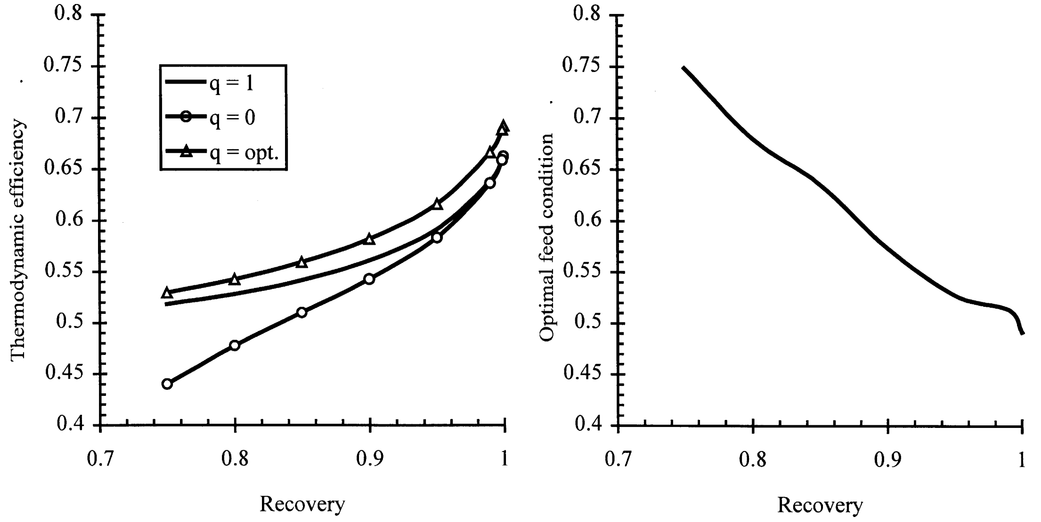
(b)

Fig. 9. Thermodynamic efficiency of a distillation column, processing feed at its optimal thermal condition, for different: (a) relative volatilities; (b) feed compositions.

thermal condition of the feed. Thermodynamic efficiency of a distillation column increases monotonically with increasing relative volatility for feed with optimal thermodynamic condition. Rate of increase in thermodynamic efficiency is very high for feed with predominantly light or heavy component, whereas, for symmetric feed (i.e., feed with nearly equal light and heavy component) rate of increase of thermodynamic efficiency is not so prominent. Therefore, it is beneficial to reduce the operating pressure and thereby increase the volatility for distillation column processing asymmetric feed. However, below a certain relative volatility, symmetric feed results in a most efficient distillation column and for a given relative volatility, thermodynamic efficiency of a distillation column is at its maximum when it processes symmetric feed.

3.3. Sharpness of separation

Thermodynamic efficiency of a distillation column changes with sharpness of separation (Fig. 10). Thermodynamic efficiency monotonically increases with increasing recovery of the components. It may be noted that the thermodynamic efficiency of a distillation column is more for two-phase feed with optimal thermal condition ($z = 0.5$



(a)

(b)

Fig. 10. Effect of product recovery on: (a) thermodynamic efficiency; (b) optimal thermal condition of the feed.

and $\alpha = 2$ for this case). Optimal thermal condition of the feed decreases with increasing recovery.

Increase in recovery calls for higher number of stages and higher utility demand. Increase in capital and operating cost are compensated by the increase in product cost due to quality improvement. Knowing the capital cost of the column, operating cost associated with utility requirement and cost of products for different quality, optimal recovery for a distillation column may be obtained.

4. Conclusions and remarks

Quantitative targets for improving energy efficiency of a distillation column may be set from the *T–H* curve. IRS curves are such a novel pair of *T–H* curves that are invariant to the column configuration (i.e., feed location in the column and number of stages) and depend on sharpness of separation and operating pressure of the column. Exergy loss in a distillation column can be graphically represented through the IRS curves by presenting them on ε–*H* plane. This paper establishes the relation between the IRS curves and exergy loss in a distillation column. Therefore, IRS curves can be utilized both for setting quantitative targets as well as exergetic efficiency of a distillation column.

Thermal condition of the feed influences the thermodynamic efficiency of a distillation column. Optimal thermal condition of the feed contributes significantly in increasing the thermodynamic efficiency of a cryogenic distillation column, since work, rather than heat, is being utilized in for sub-ambient processes such as gas separation [1,13,14]. In this paper, a general methodology has been presented to analyze the thermodynamic performance of a distillation column for different thermal condition of the feed. Dependency of thermodynamic efficiency of a distillation column on feed composition, thermal condition of the feed, relative volatility of its components and sharpness of separation have been studied in this paper. It should be noted that thermodynamic efficiency of a distillation column has been considered in this article as the sole criterion for optimizing the thermal condition of the feed. Applicability and quality of the design of a distillation column cannot possibly be judged solely on the basis of thermodynamic efficiency of the distillation column. Factors such as inefficiencies associated with converting a feed to its optimal conditions, capital and operating cost of the system, operational complexity and flexibility, etc. are also important for designing a distillation column. However, thermodynamic analysis of the distillation column does provide useful understanding and insights to the designer.

For strong non-ideal systems, there exists a tangent pinch or near-tangent pinch in either section of the column. Whenever there exists a tangent pinch or near-tangent pinch in either section of the column, the invariant rectifying curve becomes non-monotonic and shows a sharp turning point. Typical IRS curves with tangent pinch in the rectifying section is shown in Fig. 11. Vertical side of the circumscribed right-angled trapezium touches this sharp turning point (Fig. 11). This sharp turning point determines the pinch point and hence the minimum energy requirement as well as total exergy loss in the column. Unlike ideal or near ideal systems with monotonic IRS curves, intersection point of the IRS curves differs from the pinch point for non-ideal systems with tangent pinch point. However, it should be noted that the appropriate location of the feed can still be determined from the intersection point of the translated IRS curves.

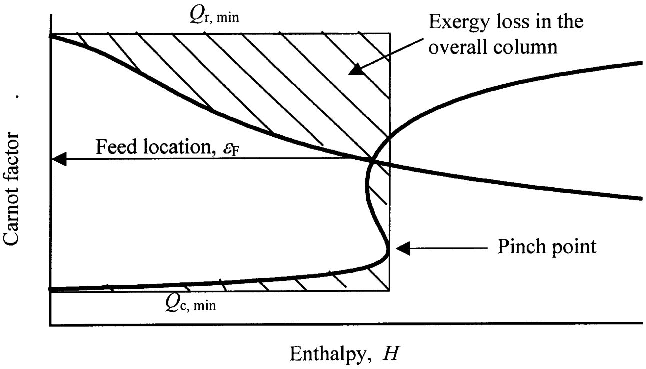


Fig. 11. IRS curves for non-ideal system showing tangent pinch at the rectifying section.

Pinch point plays a significant role in designing energy efficient heat exchanger networks [15,16]. Similarly, the significance of the pinch, in the context of distillation, may be stated as follows: no (side-)reboiling below the pinch and no (side-)condensing above the pinch. This implies that for near-ideal system with monotonic IRS curves, side-reboiler and side-condenser may be put in the stripping and the rectifying section of the column, respectively. However, for non-ideal systems with non-monotonic IRS curves, it may be possible to place side-reboiler in the rectifying section or side-condenser in the stripping section of the column. These observations are consistent with the observations of Naka et al. [17] and Agrawal and Fidkowski [18].

Furthermore, due to non-monotonicity of the IRS curves another interesting point may be observed concerning thermal condition of the feed. Whenever the feed point (i.e., the intersection point of the IRS curves) is above the pinch point, it is beneficial to preheat the feed. On the other hand, it is beneficial to precool the feed for feed point lying below the pinch point. Combining these observations with the significance of pinch stated earlier, a general principle may be stated as follows: hot utility should only be used to preheat the feed or for side-reboiling above the pinch and cold utility should only be used to precool the feed or for side-condensing below the pinch-principle equivalent to the principle employed in synthesizing energy efficient heat exchanger networks.

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