Improving Distillation Column Design Using Thermodynamic Availability Analysis

R. E. FITZMORRIS
and
R. S. H. MAH
Northwestern University
Evanston, Illinois 60201

This is a comprehensive availability analysis of conventional, vapor recompression and SRV (secondary reflux and vaporization) distillation. The physical significance of different availability losses is interpreted, and the magnitude and distribution of these losses in an ethylene-ethane separation are shown. Also discussed is how they may be reduced by design modifications.

SCOPE

In a recent study (Merix Corp. 1977) commissioned by the U.S. Energy Research and Development Administration (now Department of Energy) an audit of U.S. energy usage in distillation was prepared with reference to (a) the petroleum refinery and natural gas processing plants, and (b) the chemical industry. The combined total distillation energy consumption for 1976 was conservatively estimated to be two quads (2×10^{18}J), representing nearly 3% of our entire national energy consumption. Moreover, the audit reveals ample opportunities for energy conservation and efficiency enhancement, with payout times on typical projects ranging from 0.06 to 4.4 years (Mix et al. 1978). It was pointed out that even a 10% energy savings in distillation, easily within the realm of feasibility (Prengle et al. 1974), would be equivalent to nearly 100,000 barrels (15.9×10^6L) of petroleum per day on a nationwide basis. With such obvious incentives, it is not surprising to see a resurgence of interest in distillation column design and control (Null 1976, Mah et al. 1977, Tedder and Rudd 1978, Constellano et al. 1978, Nisenfeld 1978, Mah and Wodnik 1979, and Umeda et al. 1979).

In our previous work (Mah et al. 1977) we pointed out that distillation schemes can be evaluated on several different bases. From a practical viewpoint, an economic evaluation is the most relevant and decisive basis. But except for the most obvious comparisons, a meaningful economic evaluation cannot be carried out without the specific mechanical and equipment details and the associated cost data, which, even when available, may be highly specific to a given application.

One alternative is to make the comparison on the basis of utility requirements. Another is to evaluate the design on the basis of thermodynamics. One advantage of the last approach is that it may yield a measure of how close the design approaches the theoretical limit of an idealization. In so doing, it provides a guide for further design improvements. It should be pointed out, however, that the earlier characterization of distillation efficiency (Freshwater 1951) with reference to an idealized reversible isothermal model is inappropriate for most distillation schemes in which temperature gradient is a key feature of the underlying process. A more recent paper by Umeda et al. (1979) focuses on heat transfer irreversibilities, but appears to ignore other irreversibilities in distillation. In spite of its obvious importance, a thorough examination of distillation from the viewpoint of thermodynamic efficiency has not been previously reported.

In this article, we present such an analysis and show how it can impart useful understanding and insights, which will lead to better designs of distillation systems. We apply this analysis to a specific separation, the C_2 (ethylene-ethane) splitter, and report the numerical results of computer simulation.

CONCLUSIONS AND SIGNIFICANCE

The availability losses in a distillation system are shown to be separately attributable to heat transfer, compression, expansion and stage mixing, their effects on net work requirement being additive. The last of these terms accounts for irreversibilities due to the mixing of streams with non-equilibrium compositions and temperatures, which occurs on all stages. The magnitude of this term is exactly equal to the difference between the Carnot cycle work required to transfer heat to and from the environment for real and reversible distillation. For adiabatic distillation at minimum reflux, the heat duties are exactly the same as for isobaric reversible distillation.

In general, availability losses may be classified into those arising out of inefficient coupling of heat sources and sinks internal to the columns and those arising out of inefficient exchanges of heat and work with the environment. By lowering external reflux ratio and increasing the number of stages, the use of secondary reflux and vaporization (SRV) can substantially reduce both stage mixing irreversibilities and net work requirements. Several thermodynamic efficiencies for measuring internal and external irreversibilities are defined and may be used as diagnostic aids in design improvements.

THERMODYNAMIC AVAILABILITY CHANGES IN DISTILLATION

The general distillation system to be considered can undergo material, heat and work exchanges with the environment. Let m_i and h_i be the molal flow rate and enthalpy of an input or output stream i to the system, let q_j be the rate of heat exchange with the environment through an auxiliary heat reservoir maintained at temperature T_j, and let w_l be the work done by a part of the system. Then the first law of thermodynamics stipulates that

$$\frac{d(MU)_{\text{syst}}}{dt} = \sum (mh)_i + \sum q_j - \sum w_l \qquad (1)$$

where M_{syst} is the mass of the system, U_{syst} is the internal energy averaged over the system and the accumulation term reflects the usual assumption that kinetic and potential energy contributions

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Address correspondence to R. S. H. Mah

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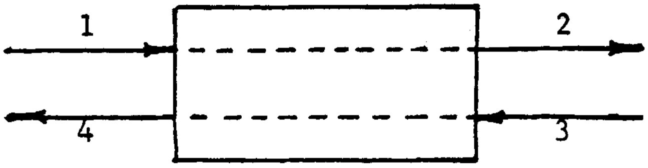

Figure 1. A two-stream heat exchanger.

may be neglected. The summations are to be carried out under the sign convention that heat and material inputs to the system and work done by the system are taken as positive. For the same system the second law of thermodynamics stipulates the following inequality

$$\frac{d(MS)_{\text{syst}}}{dt} \geqslant \sum(mS)_i + \sum q_j/T_j \tag{2}$$

where S_{syst} denotes the entropy averaged over the system and S_i is the entropy of the material in stream i.

Note that as an inequality constraint in design optimization, Relation (2) is not very useful, because it can never be active or binding in a real situation—the equality will hold only for a reversible process. But, it can be readily converted into an equation by introducing an additional term σ to account for the rate or irreversible entropy increase, as follows

$$\frac{d(MS)_{\text{syst}}}{dt} = \sum(mS)_i + \sum q_j/T_j + \sigma \tag{3}$$

Denbigh (1956) attributed the innovative use of this term to de Donder. Equation (3), a restatement of the second law, is sometimes referred to as an "entropy balance." But, unlike Equation (1), it is not a conservation statement, since entropy is clearly a non-conserving quantity. In fact, since all its other terms may be measured or computed, Equation (2) may be viewed as the definition of σ. The viewpoint turns out to be very useful in our later discussion.

Also, instead of stating the second law in terms of entropy, as we did in Equation (3), it will be very much more useful to restate it in terms of work. Such a statement is obtained by multiplying Equation (3) by the temperature of the environment T_o and subtracting the result from Equation (1), as

$$\frac{d}{dt}\,[M(U - T_oS)]_{\text{syst}} = \sum\,[m(h - T_oS)]_i + \sum\left(1 - \frac{T_o}{T_j}\right) q_j - \sum w_l - T_o\sigma \tag{4}$$

We emphasize the distinction between the environment state and the reference state used in the computation of a thermodynamic property. The former is the prevailing state of the infinite reservoir. Thus, T_o is usually in the range of 280-300 K. But the latter may differ for different properties. For processes involving chemical reactions, de Nevers and Seader (1979) choose the elements as gases at 0 K and zero pressure as the enthalpy reference state and the compounds as ideal gases at 255.37 K at unit pressure as the entropy reference state. For our computation the reference state is immaterial. Since no chemical reactions are involved, only ideal gas heat capacity correlations (ASHRAE Handbook 1977, Angus et al. 1974) are required.

The derivation of Equation (4) follows similar developments presented earlier by Denbigh (1956), de Nevers and Seader (1979), and others, who have also given physical interpretations to its various terms. Denbigh (1956) showed that the term $(1 - T_o/T_j)q_j$ is the Carnot cycle work required to restore q_j to the heat reservoir at T_j at the expense of the environment. The quantity $(h - T_oS)$ is the widely accepted steady flow availability function B (see, for example, Haywood 1974). De Nevers and Seader (1979) gave an interpretation of $T_o\sigma$ as the sum of (mechanical) lost work and the Carnot cycle work needed to restore the irreversible conversion of work to heat. In real processes, the irreversible creation of entropy (σ) or the "waste work" ($T_o\sigma$) can arise from numerous sources such as finite temperature driving force in heat transfer, mass transfer by mixing, compression, expansion and, of course, chemical reaction. In the ensuing development, we attempt to identify and interpret those losses of availability or waste work associated with the distillation processes. Since we are only concerned with the steady state process, the L.H.S. (left hand side) of Equation (4) will be taken as zero in all subsequent analysis.

Availability Loss Through an Expansion Valve

If we make the usual simplification that the expansion is adiabatic, Equation (4) reduces to

$$(mB)_{\text{in}} - (mB)_{\text{out}} = (T_o\sigma)_{\text{valve}} = \beta_v \tag{5}$$

The availability loss can therefore be computed, if we know the flow rate and conditions of the inlet and the outlet streams of the valve. We introduce an abbreviated notation β_v to designate this quantity.

Availability Loss in Heat Exchange

For simplicity, we consider a two-stream heat exchanger involving only two temperature levels, T_H and T_C. With reference to Figure 1 (since there is no heat exchange with any auxiliary heat reservoir), we can again simplify Equation (4) to

$$(T_o\sigma)_{\text{heat exch}} = \sum_{i=1}^{4}\,[m(h - T_oS)]_i$$

$$= \sum_{i=1}^{4}(mh)_i - \sum_{i=1}^{4}(mT_oS)_i \tag{6}$$

Enthalpy balance demands that the first term on the right-hand side be zero. Hence, if q is the total rate of heat transfer, we have

$$\beta_{hx} = (T_o\sigma)_{\text{heat exch}}$$

$$= T_o[m_1(S_2 - S_1) + m_2(S_4 - S_3)]$$

$$= T_o(q/T_C - q/T_H) \tag{7}$$

Equation (7) may be restated and given the following interpretations for different temperature distributions

(a) For $T_H > T_C > T_o$

$$\beta_{hx} = q(1 - T_oT_H) - q(1 - T_oT_C) \tag{7a}$$

The first term represents the Carnot cycle work required to transfer heat q over unit time to the hot stream T_H from the environment T_o. The second term with the minus sign represents the work derived from a Carnot cycle operating between T_C and T_o to remove heat q over unit time from the cold stream at T_C.

(b) For $T_o > T_H > T_C$

$$\beta_{hx} = (qT_o/T_C)(1 - T_C/T_o) - (qT_o/T_H)(1 - T_H/T_o) \tag{7b}$$

The first term represents the Carnot cycle work required to transfer heat q over unit time from the cold stream at T_C to the environment T_o. The second term with the minus sign represents the work derived from a Carnot cycle operating between T_o and T_H to transfer heat q over unit time to the hot stream at T_H.

(c) For $T_H > T_o > T_C$

$$\beta_{hx} = (qT_oT_C)(1 - T_CT_o) + q(1 - T_oT_H) \tag{7c}$$

In this case, both terms represent work required for Carnot cycles operating between T_o and T_C and between T_H and T_o transferring heat q over unit time from T_C and to T_H.

For heat exchange over continuous ranges of temperatures, Equation (7) may be replaced by

$$\beta_{hx} = T_o\int(dqT_C) - T_o\int(dqT_H) \tag{8}$$

and the same interpretations apply.

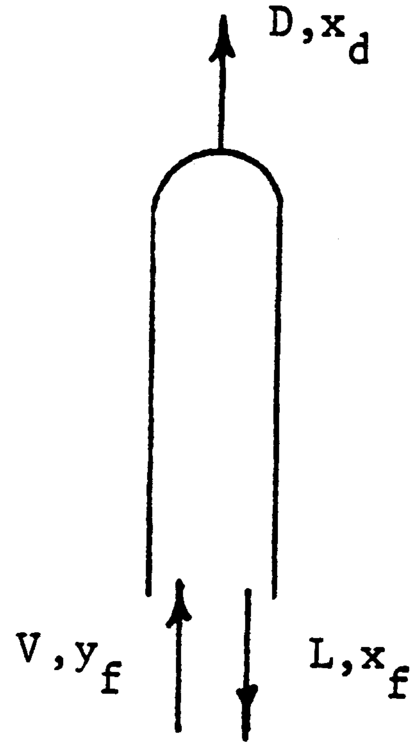

Figure 2. Material flows in a rectifying section.

Heat Duties in Isobaric Reversible Distillation and Adiabatic Distillation at Minimum Reflux

Consider binary distillation with specified feed, distillate and bottoms compositions and flow rates. We assume the column operates at a specified pressure, with zero pressure drop through its stages. Let Q_R be the cumulative rate of heat removal from its rectifying section, including the condenser, and let Q_S be the cumulative rate of heat addition to its stripping section, including the reboiler. Then for isobaric reversible distillation (Mah and Wodnik 1979), as for adiabatic distillation at minimum reflux, we have, for the rectifying section shown in Figure 2,

$$Vy_f = Lx_f + Dx_d \tag{9}$$

$$V = L + D \tag{10}$$

$$Q_R = Dh_d + Lh_L - VH_V \tag{11}$$

where V and L are the vapor and liquid flow rates entering and leaving the rectifying section, y_f and x_f the component mole fractions in these streams, H_V and h_L the corresponding enthalpies, and D, x_d and h_d are the flow rate, component mole fraction and enthalpy associated with the distillate. Since L, V and Q_R are the only three unspecified variables in these three equations, we conclude that their values must be the same for both isobaric reversible distillation and for adiabatic distillation at minimum reflux. A similar set of equations can be written for the stripping section with the overall conclusion that the Q_R and Q_S for isobaric reversible distillation and for adiabatic distillation at minimum reflux are the same.

Now, if the desired separation is carried out reversibly, Equation (4) stipulates that the minimum work required is

$$\begin{aligned} w_{\min} &= -\sum w_i = -\sum (mB)_i \\ &= (mB)_{\text{distillate}} + (mB)_{\text{bottoms}} - (mB)_{\text{feed}} \end{aligned} \tag{12}$$

which, in the case of isobaric reversible distillation with continuous heat exchange with the environment throughout the height of the column, is

$$w_{\min} = \int_0^{Q_S} (1 - T_o/T_S)dq - \int_0^{Q_R} (1 - T_o/T_R)dq \tag{13}$$

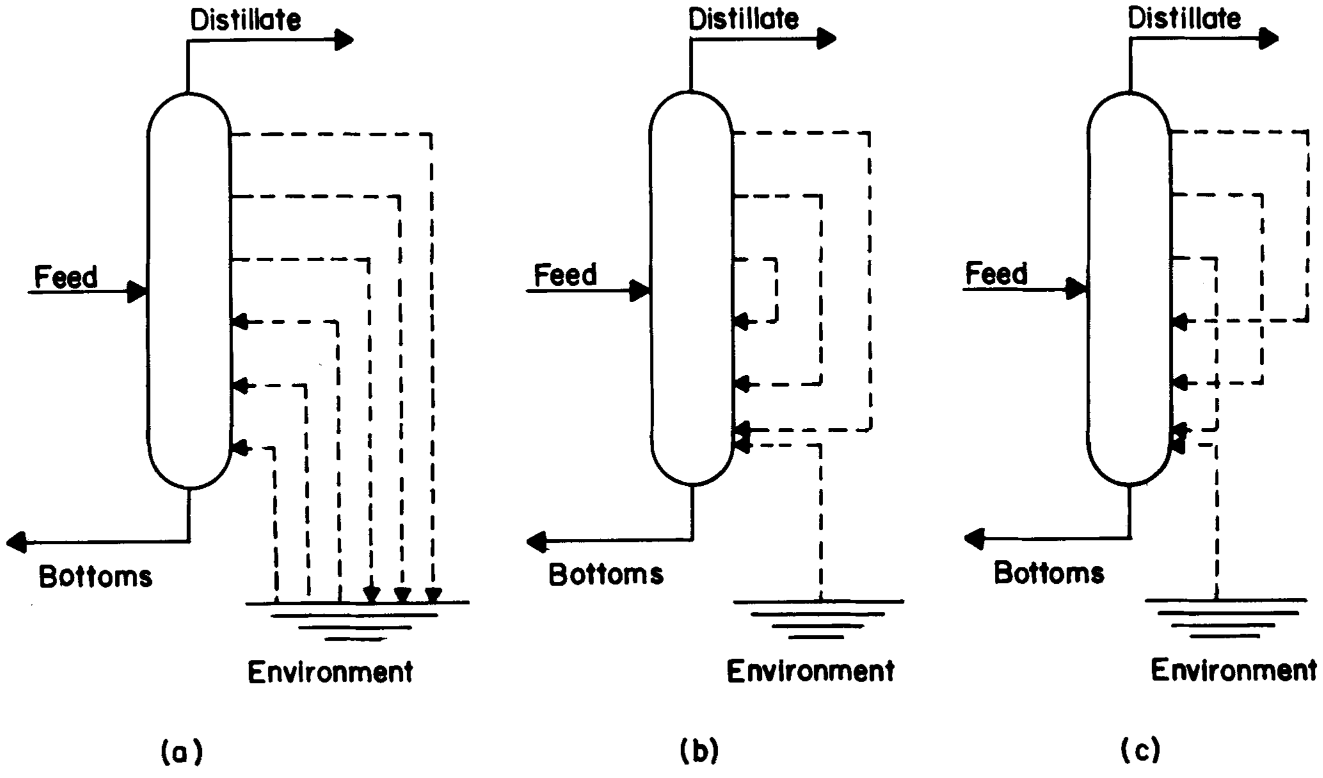

Figure 3. Models of isobaric reversible distillation.

A further stipulation is imposed by the overall enthalpy balance

$$F\,H_f + Q_S = Wh_b + Dh_d + Q_R \tag{14}$$

where F and H_f are the molal flow rate and enthalpy of the feed, and W and h_b are the molal flow rate and enthalpy of the bottoms. Equations (12), (13) and (14) define $w_{\min}$, Q_R and Q_S for isobaric reversible distillation and adiabatic distillation at minimum reflux.

Notice that Equation (13), which corresponds to the scheme shown in Figure 3a, is not the only form of reversible distillation for the desired separation. If we couple the heat sources at T_R with the heat sinks at T_S by reversible heat pumps, and supply or remove the residual heat duty by reversible heat pumps or expansion engines at the expense of the environment, we arrive at another model of reversible distillation which will require the same $w_{\min}$ for the separation. Thus, if

$$Q^* = Q_R + \int_0^{Q^*} (1 - T_R/T_S)dq \le Q_S \tag{15}$$

then

$$w_{\min} = \int_0^{Q^*} (1 - T_R/T_S)dq + \int_{Q^*}^{Q_S} (1 - T_o/T_S)dq \tag{16}$$

Two such schemes are shown in Figures 3b and 3c.

Availability Losses in Distillation Stages

Consider first the conventional adiabatic distillation which involves no shaft work, compression or expansion. The only heat exchanges with the environment are through the condenser at T_D and the reboiler at T_B. Let us exclude any heat transfer irreversibilities by defining those parts of the system boundary to be at temperatures T_D and T_B, respectively. Applying Equation (4) to such a system we obtain

$$\sum (mB)_i + (1 - T_o/T_B)Q_S - (1 - T_o/T_D)Q_R = (T_o\sigma)_{\text{stage mixing}} = \beta_{sm}$$

or

$$\beta_{sm} = (1 - T_o/T_B)Q_S - (1 - T_o/T_D)Q_R - w_{\min} \tag{17}$$

Note that β_{sm} accounts for all the irreversibilities which occur on all stages: the mixing of streams at non-equilibrium temperatures and compositions. Since there is no other irreversibility associated with the system, β_{sm} is also the sum of all $T_o\sigma$ terms when Equation (4) is applied stage by stage to the whole column.

In the special case of minimum reflux, the heat duties Q_R and Q_S are the same as those for the reversible distillation. Equations (13) and (17) may be combined to give

$$\beta_{sm} = (1 - T_o/T_B)Q_S - (1 - T_o/T_D)Q_R - \int_0^{Q_S} (1 - T_o/T_S)dq + \int_0^{Q_R} (1 - T_o/T_R)dq \tag{18}$$

The first two terms represent the net work required to transfer heat Q_R from the condenser at T_D to the environment at T_o and to transfer heat Q_S to the reboiler at T_B from the environment at T_o. The last two terms represent the net work required to transfer heat Q_R from the rectifying section differentially over its temperature range to the environment at T_o and to transfer heat Q_S to the stripping section differentially over its temperature range from the environment at T_o. It is interesting to note that the difference between these two net work terms measures exactly the cumulative availability losses on all the stages. The significance of this physical interpretation is that any design modification which reduces this net work difference will also reduce availability losses due to stage mixing, and vice versa.

A particular form of the reversible distillation is described by Equations (15) and (16). By combining Equation (16) with its counterpart to Equation (17) we obtain

$$\beta_{sm} = -\left\{(1 - T_B/T_D)Q_R + \int_0^{Q^*} (1 - T_R/T_S)dq\right\} + \left\{(1 - T_o/T_B)\Delta Q_S - \int_{Q^*}^{Q_S} (1 - T_o/T_S)dq\right\} \tag{19}$$

where

$$\Delta Q_S = Q_S - Q_R\,T_B/T_D \ge 0 \tag{20}$$

Equation (19) shows that the reduction of availability losses can be conveniently grouped into two terms: The first term is associated with the internal coupling of heat sources and heat sinks and the second term is associated with heat exchanges with the environment. The magnitude of the first term gives a measure of maximum potential improvement which might be achieved through enhanced internal heat integration. On the other hand, the second term is a measure of the importance of energy integration with the environment. This term will dominate, if $|T_R - T_o| >> |T_S - T_R|$.

Finally, it should be noted that for conventional distillation under realistic conditions, the availability losses due to stage mixing will be greater than that given by Equation (18). This is so because at non-minimal reflux, Q_S and Q_R will both be larger by approximately the same amount.

Thermodynamic Efficiencies

To summarize the discussion so far we may restate Equation (4) for steady state distillation as

$$\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i + \sum_{\substack{\text{process}\\\text{utilities}}} (mB)_k + \sum_{\substack{\text{heat}\\\text{sources}\\\text{and sinks}}} (1 - T_o/T_j)q_j - \sum w_l - \beta_v - \beta_{hx} - \beta_{sm} = 0 \tag{21}$$

It is convenient to break up the availability flows into two terms: The first term accounts for the feeds and products. The second accounts for process utilities, i.e. process streams which exchange heat, but not material, with the system. The process utility term could be positive or negative, depending on whether there is a net availability input or output. Strictly speaking, we should also include in Equation (21) a term for availability losses due to compression. But in our simulation, isentropic compression was assumed for simplification.

Based on Equation (21), we can define thermodynamic efficiency in several ways depending on our viewpoints. Suppose that $w_{\min}$ is positive, i.e., work must be done on the system to obtain the desired separation, as is usually the case. Then viewing $\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i$ as the desired result we may define

$$\eta_1 = -\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i \Big/ \left(-\sum w_l\right) \tag{22}$$

or

$$\eta_2 = -\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i \Big/ \left[-\sum w_l + \sum_{\substack{\text{process}\\\text{utilities}}} (mB)_k\right] \tag{23}$$

On the other hand, we may view

$$\left[\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i + \sum_{\substack{\text{process}\\\text{utilities}}} (mB)_k\right]$$

as the desired result and define

$$\eta_3 = -\left[\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i + \sum_{\substack{\text{process}\\\text{utilities}}} (mB)_k\right] \Big/ \left(-\sum w_l\right) \tag{24}$$

Finally, we may restrict the considerations to the column itself (excluding the valves and the heat exchangers) and define

$$\eta_4 = -\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i \Big/ \left[-\sum_{\substack{\text{process}\\\text{streams}}} (mB)_i + \beta_{sm}\right] \tag{25}$$

When the process utility term is absent, $\eta_3 = \eta_2 = \eta_1$ and the definitions become identical to that given by de Nevers and

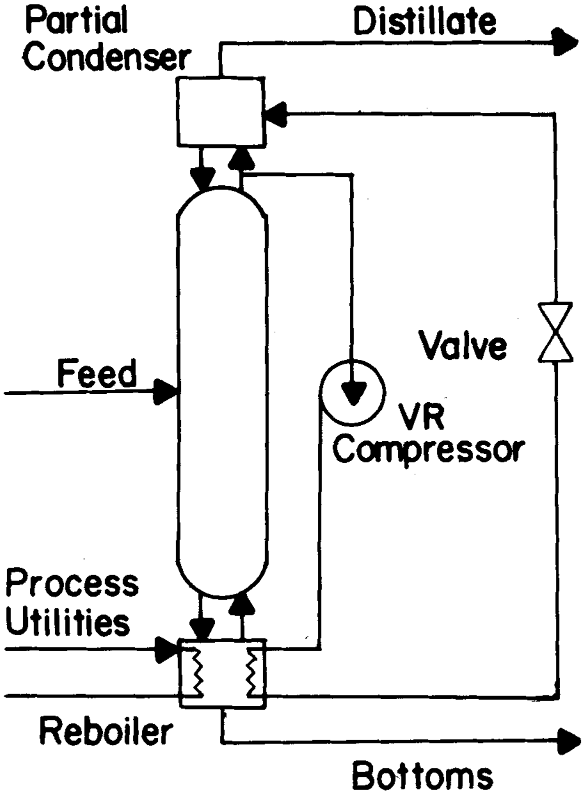

Figure 4. Vapor recompression distillation.

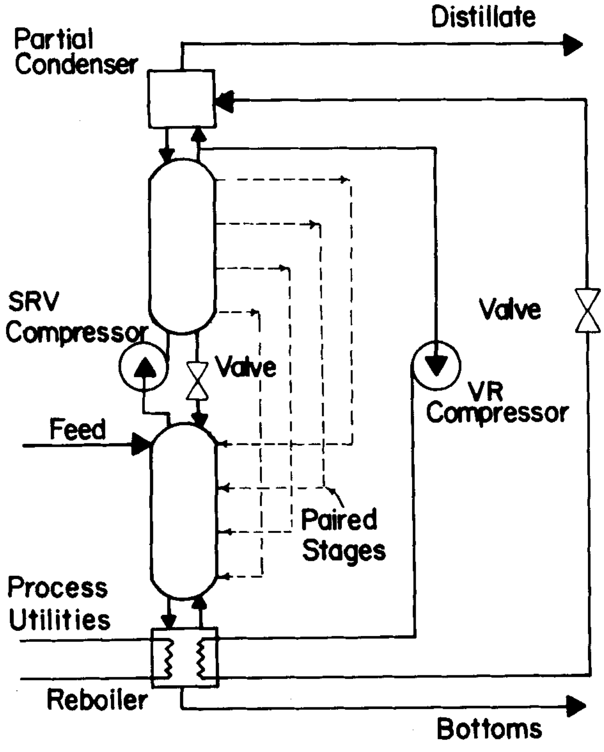

Figure 5. SRV distillation.

Seader (1979, Eq. 49). For work producing system, the efficiencies would be the reciprocals of those defined above, i.e. $\eta_i' = 1/\eta_i$.

SIMULATION STUDIES

To illustrate the utility of availability analysis in design studies, we apply this approach to evaluate the alternative designs for a C_2 (ethylene-ethane) splitter. This separation is normally effected under pressure with the utility cost dominated by the compression requirements, and the capital investment similarly dominated by the compressor cost. Any reduction in power consumption is likely to reduce both the operating and capital costs. In this investigation, we attempt to discover both the magnitude and the distribution of availability losses in the various distillation systems, and how they may be modified by a change in design and operating conditions. Three types of designs are considered: The conventional distillation, the vapor recompression distillation and the SRV distillation with vapor recompression. The last two are shown schematically in Figures 4 and 5.

As in our previous investigation (Mah et al. 1977), the Soave-Redlich-Kwong equation of state was used to estimate the thermophysical properties. A modified Wang-Henke method was used to carry out the distillation calculations. The simulation was straightforward, except for the SRV distillation at low reflux ratios, for which the convergence became quite sensitive to the quality of initial guesses.

A procedure which successively reduces the reflux ratio as it increases heat transfer areas was found to work satisfactorily. In all simulation runs, theoretical stages with zero pressure drops were assumed. The compression was assumed to be isentropic, and the expansion through valves was assumed to be isenthalpic. An overall heat transfer coefficient of 567.8 J/m^2·s·K was assumed. These simplifications are not expected to affect our analysis materially. In fact, with a little more computation, they can be readily eliminated.

It is customary to classify the C_2 splitters according to their operating pressures. The "high pressure" columns operate at around 1.5×10^6 N/m^2; the "low pressure" columns operate between 1.5×10^5 and 6×10^5 N/m^2 (Ruhemann and Charlesworth 1966). For our conventional column, we picked an operating pressure of 1.7×10^6 N/m^2. The feed conditions and performance specifications for the conventional distillation run are the same as for all subsequent runs, except for the pressure and the thermal conditions of the feed, shown in Table 1.

TABLE 1. FEED CONDITIONS AND PERFORMANCE SPECIFICATIONS

Feed Conditions:	50% Ethylene	
	50% Ethane	
	Pressure = 304,000 N/m^2	(VR and SRV)
	1,723,000 N/m^2	(conventional)
	Temperature = 199 K	
	Flow Rate = 0.126 kg-mole/s.	
	Fraction Liquefied = 0.71	(VR and SRV)
	Saturated Liquid	(conventional)
Distillate Specification:	99 ± 0.05% Ethylene	
	Pressure = 557,300 N/m^2	(VR and SRV)
	1,723,000 N/m^2	(conventional)
	Flow Rate = 0.0630 kg-mole/s	
Condenser Type:	Partial	

	Conventional	Vapor recompression	SRV
Column Pressure, N/m^2			
Rectifier	1,723,000	304,000	304,000
Stripper	1,723,000	304,000	557,300
Distillate			
Temperature, K	239	191	206
Bottoms			
Temperature, K	260	208	208

TABLE 2. AVAILABILITY ANALYSIS OF CONVENTIONAL DISTILLATION

Case	CONV
Number of Stages	44
Reflux Rate	4.9
Heat Transfer Rate, W	10,932,000
Net Work Input, W	879,706
Refrigeration	1,245,366
Process Utilities	−365,660
Minimum Work, W	51,874
Availability Loss, W	827,950
Refrigeration Heat Transfer	263,422
Valve	284,344
Reboiler Heat Transfer	124,046
Stage Mixing	156,138

DISCUSSION OF RESULTS

Tables 2-6 show the results of computer simulation with breakdown of work input and availability losses as shown in Equation (21). The item labeled "distillate compressor" is the work needed to bring the distillate vapor to the required pressure of 5.57×10^5 N/m^2. The negative value for process utilities indicates the work which may be obtained by a Carnot engine operating between the environment temperature T_o and the temperature of the process utilities. Since each item of work input and availability losses was separately computed from an availability balance, the discrepancy between the net work input and the sum of minimum work required and total availability loss is a measure of computational errors. The magnitude of these errors is determined by the convergence conditions set for the iterative calculations.

As one might expect, the magnitude of β_{hx} depends critically on the temperature approach (driving force), which varies in different parts of the system. To compare β_{hx} for different designs, each with a different temperature profile and heat load distribution, we choose to make the availability loss per unit of heat transferred the same for all runs, by adjusting the appropriate (refrigerant, vapor recompression or rectifying column) pressure. This value was held to 0.034 ± 0.00005, which corresponds to an average temperature approach of about 5 K. An examination of the distribution of availability losses in Tables 3-6 reveals that in all cases, β_{sm} accounts for the largest single loss under the temperature approach condition that we have chosen.

It should be evident from the results that the conventional "high pressure" system requires about three times as much in net work input at the vapor recompression or the SRV distillation. Much of the availability losses occurs in the refrigeration system. These losses can be eliminated by vapor recompression or SRV distillation. Therefore, in the subsequent runs, the comparison of vapor recompression and SRV distillation was emphasized.

Table 3 compares SRV distillation runs with the same number of stages, but different external reflux ratios (1.95 to 3.20). The run SRVH1 with zero interstage heat transfer is, effectively, the vapor recompression distillation with the rectifying section operating at a higher pressure than the stripping section. As the external reflux ratio is reduced, the interstage heat transfer must be increased to meet the same performance specifications. But contrary to first expectations, the total availability loss is little affected by this redistribution of heat load. As the interstage heat transfer increases, β_{sm} decreases. But the decrease is almost exactly compensated by the increase in β_{hx}, due to the larger total (stripping section) heat transfer rate. The latter is defined as the sum of heat transfer rates across all heat transfer surfaces contained in the system. It is tabulated in column 4 of Table 7

Table 4 shows the simulation results for vapor recompression distillation, with different numbers of stages and different reflux ratios for meeting the same performance specifications. Tables 5 and 6 show the simulation results for the corresponding SRV distillation runs. Because the same separation can be obtained with different reflux ratios for the same number of stages in SRV distillation (see, for example, Table 3), it is necessary to specify an additional condition to define the runs uniquely. We chose to specify the total heat duty (condenser and interstage heat transfer) of the rectifying section. Table 5 contains the runs with the same "high" heat duties and Table 6 contains the runs with the same "low" heat duties. An examination of Tables 4, 5 and 6 shows that the net work input may be reduced by reducing the external reflux ratio and correspondingly increasing the number of stages. For vapor recompression distillation, the minimum is attained at the minimum reflux ratio estimated to be 2.16, when Q_R and Q_S would be the same as for the isobaric reversible distillation. A similar minimum exists for the SRV distillation but the net work input is generally significantly lower. Compare, for example, runs VR1 and SRVL3. Comparison of runs VR2 and SRVH4 reveals that for the same stripping section heat transfer rate but different numbers of stages the SRV distillation requires 22% less net work input and gives rise to a β_{sm}, which is 50% lower than the vapor recompression distillation.

To a good first approximation, the compressor work input and the availability losses are all proportional to Q_R or Q_S, which are linear functions of reflux ratio. The linearity is verified by the plots shown in Figures 6, 7 and 8.

Finally, the thermodynamic efficiencies for all the simulation runs are computed and tabulated in Table 7. The three types of thermodynamic efficiencies measure different aspects of system performance. η_4 is a measure of the systems performance as a separation device excluding any consideration of heat transfer, expansion and other physical steps. On this basis, the SRV distillation outperforms the other designs by a factor of two. η_1 is

TABLE 3. AVAILABILITY ANALYSIS OF SRV DISTILLATION AT CONSTANT NUMBERS OF STAGES

Case			SRVH1	SRVC1	SRVC2	SRVC3
Number of Stages			30	30	30	30
Reflux Ratio			3.20	2.35	2.15	1.95
Heat Transfer Area, m^2/stage			0	24.15	30.19	39.48
Work Input, W	SRV Compressor		241,206	262,454	266,730	275,859
	VR Compressor		90,289	65,737	60,255	56,355
	Distillate Compressor		0	0	0	0
	Process Utilities		−8,934	−10,199	−10,656	−8,565
	Net Work Input		322,561	317,992	316,329	323,648
Minimum Work, W			67,524	66,146	65,702	65,882
Availability Losses, W	Heat Transfer	*Interstage*	0	32,222	39,102	46,329
		Condenser/Reboiler	87,049	63,760	58,812	55,335
	Valve	Interstage	14,801	16,530	16,908	17,652
		Condenser/Reboiler	5,760	4,175	3,830	3,641
	Stage Mixing		147,877	140,274	135,266	134,489
	Total Availability Loss		255,487	256,961	253,918	257,446

TABLE 4. AVAILABILITY ANALYSIS OF VAPOR RECOMPRESSION DISTILLATION

Case			VR1	VR2	VR3	VR4
Number of Stages			30	28	26	24
Reflux Ratio			2.63	2.75	2.88	3.10
Stripping Section Heat Transfer Duty, W			2,465,666	2,566,834	2,668,588	2,851,435
Work Input, W	VR Compressor		260,409	272,784	284,654	306,747
	Distillate Compressor		60,801	60,801	60,801	60,801
	Process Utilities		−28,670	−23,911	−17,593	−10,898
	Net Work Input		292,540	309,674	327,862	356,650
Minimum Work, W			51,773	51,451	50,162	49,927
Availability Losses, W	Heat Transfer	Interstage	0	0	0	0
		Condenser/Reboiler	83,911	87,273	90,782	97,077
	Valve	Interstage	0	0	0	0
		Condenser/Reboiler	28,727	30,091	31,357	33,772
	Stage Mixing		125,527	141,884	154,969	177,278
	Total Availability Loss		238,165	259,248	277,108	308,127

TABLE 5. AVAILABILITY ANALYSIS OF SRV DISTILLATION, RECTIFYING SECTION DUTY = − 2,432,490 W

Case			SRVH1	SRVH2	SRVH3	SRVH4
Number of Stages			30	34	40	46
Reflux Ratio			3.2	2.25	1.6	1.17
Heat Transfer Area, m²/stage			0.0	14.86	22.65	25.55
Work Input, W	SRV Compressor		241,206	241,845	242,760	241,584
	VR Compressor		90,289	61,023	43,266	32,472
	Distillate Compressor		0	0	0	0
	Process Utilities		−8,934	−19,979	−26,475	−31,017
	Net Work Input		322,561	282,889	259,551	243,039
Minimum Work, W			67,524	66,351	67,494	67,377
Availability Losses, W	Heat Transfer	Interstage	0	27,939	44,659	53,988
		Condenser/Reboiler	87,049	59,488	42,642	33,286
	Valve	Interstage	14,801	14,853	14,919	14,825
		Condenser/Reboiler	5,760	3,815	2,702	2,054
	Stage Mixing		147,877	110,904	87,248	71,580
	Total Availability Loss		255,487	216,999	192,170	175,733

a measure of overall system performance without taking any credit of potential work recovery. In other words, it measures the performance of a distillation system—in isolation from the rest of the process. Using this measure, the SRV distillation is still substantially more efficient than the vapor recompression distillation with the conventional distillation decidedly inferior to both. By contrast, η_3 which takes full credit of work recovery can be thought of as a measure of overall performance of a distillation system which is fully integrated with the other parts of the process. The striking difference between η_3 and η_1 for the conventional distillation reflects the dominant effect of "process utilities." These three efficiencies stake out the extremes with the reality somewhere in between.

CLOSING REMARKS

Our investigation shows that in a system as complex as distillation, there is no unique definition of thermodynamic efficiency. Of the various definitions presented earlier, the three measures η_1, η_3 and η_4 seem to be most useful. They stake out three limiting cases of system performance. Which one of these is most appropriate depends on the exact context and specific circumstances of the application.

For design purposes, the irreversibilities associated with fractional distillation may be conveniently classified into two broad categories: Irreversibilities associated with the concentration and temperature differences internal to the columns, and irreversibilities arising out of material and energy exchanges between the system and its environment. The former can be

TABLE 6. AVAILABILITY ANALYSIS OF SRV DISTILLATION, RECTIFYING SECTION DUTY = − 2,236,425 W

Case			SRVL1	SRVL2	SRVL3
Number of Stages			36	40	46
Reflux Ratio			2.91	2.33	1.87
Heat Transfer Area, m²/stage			0	6.5	10.13
Work Input, W	SRV Compressor		224,894	225,017	225,293
	VR Compressor		82,676	61,055	46,924
	Distillate Compressor		0	0	0
	Process Utilities		-18,094	−26,560	−32,076
	Net Work Input		289,476	259,512	240,141
Minimum Work, W			69,194	66,674	66,820
Availability Losses, W	Heat Transfer	Interstage	0	21,341	35,208
		Condenser/Reboiler	80,230	59,049	45,214
	Valve	Interstage	13,460	13,460	13,481
		Condenser/Reboiler	5,291	3,755	2,818
	Stage Mixing		122,704	95,724	76,922
	Total Availability Loss		221,685	193,329	173,644

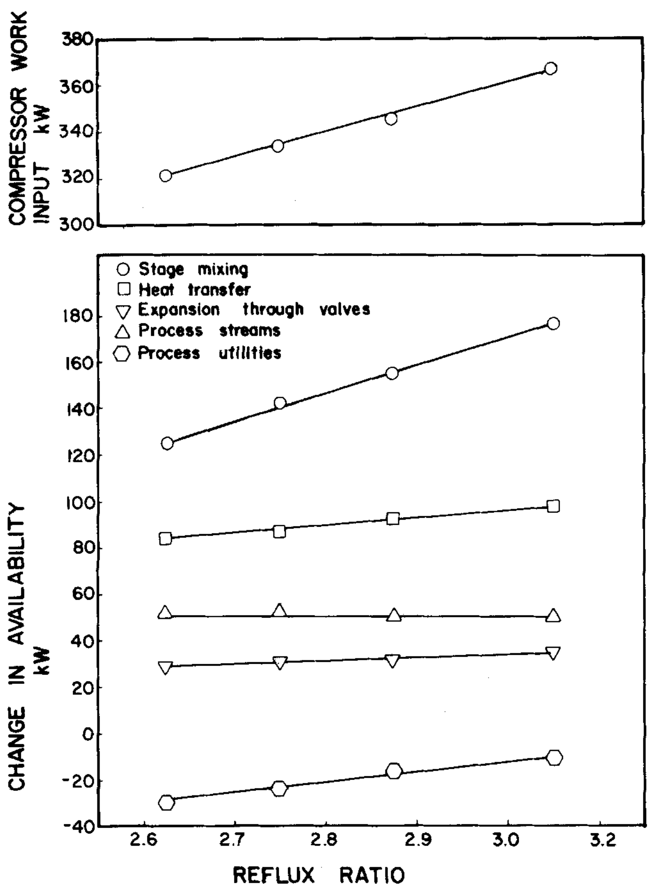

Figure 6. Availability changes in vapor recompression distillation.

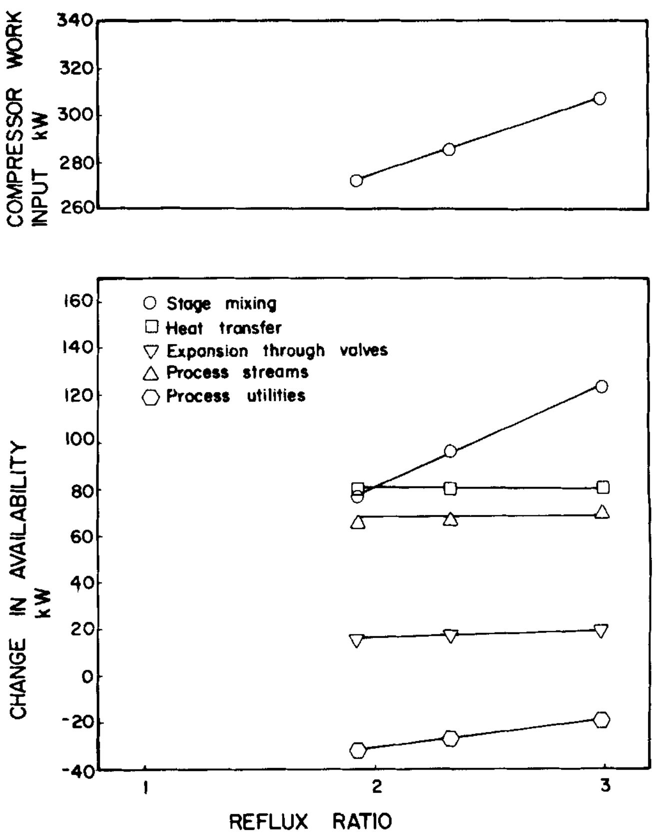

Figure 7. Availability changes in SRV distillation. Rectifier heat duty = 2.36 MW.

thought of as the departure from the reversible coupling of energy sources and sinks—the first term on the R.H.S. of Equation (19) (or as β_{sm} which is very nearly the same thing). The latter includes irreversibilities associated with external work and heat transfer, compression and expansion. A large value of β_{sm} or a small value of η_4 is an indication that the "internal" irreversibilities are significant. They may be substantially reduced by lowering the external reflux ratio and increasing the number of stages in the SRV distillation scheme. The exact extent of reduction desired will, of course, depend on the economic balance between capital investment and operating costs. But our simulation indicates that a substantial efficiency enhancement may be obtained with only a modest increase in the number of stages. On the other hand, if the situation is dominated by "external" irreversibilities as manifested by large disparities between η_3 and η_1 or large values of the "process utilities" term, it may repay well to improve the external process integration. A strategy for the design improvement in this direction was presented recently by Umeda et al. (1979).

TABLE 7. WORK REQUIREMENTS AND EFFICIENCIES

Case	Reflux Ratio	Number of Stages	Total Heat Transfer Rate* MW	Compression Work Required W	Net Work Required W	Efficiency Column η_4	Efficiency System η_1	Efficiency System η_3
CONV	4.90	44	10.93	1,245,366	879,706	0.249	0.042	0.335
VR1	2.63	30	2.47	321,210	292,540	0.292	0.202	0.253
VR2	2.75	28	2.57	333,585	309,674	0.266	0.154	0.225
VR3	2.88	26	2.67	345,455	327,862	0.245	0.146	0.197
VR4	3.10	24	2.85	367,548	356,650	0.220	0.135	0.165
SRVH1	3.20	30	2.57	331,495	322,561	0.313	0.203	0.230
SRVH2	2.25	34	2.57	302,868	282,889	0.374	0.219	0.285
SRVH3	1.60	40	2.57	286,026	259,551	0.436	0.236	0.328
SRVH4	1.17	46	2.57	274,056	243,039	0.485	0.246	0.359
SRVL1	2.91	36	2.36	307,570	289,476	0.361	0.224	0.283
SRVL2	2.33	40	2.36	286,072	259,512	0.411	0.233	0.325
SRVL3	1.90	46	2.36	272,217	240,141	0.465	0.245	0.363
SRVC1	2.35	30	2.82	328,191	317,992	0.320	0.201	0.233
SRVC2	2.15	30	2.89	326,985	316,329	0.327	0.199	0.231
SRVC3	1.95	30	2.99	332,214	323,648	0.329	0.199	0.224

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* The total heat transfer duty is the same as the stripping section heat transfer duty for all except the conventional distillation for which it is the sum of column duties and refrigeration condenser duties.

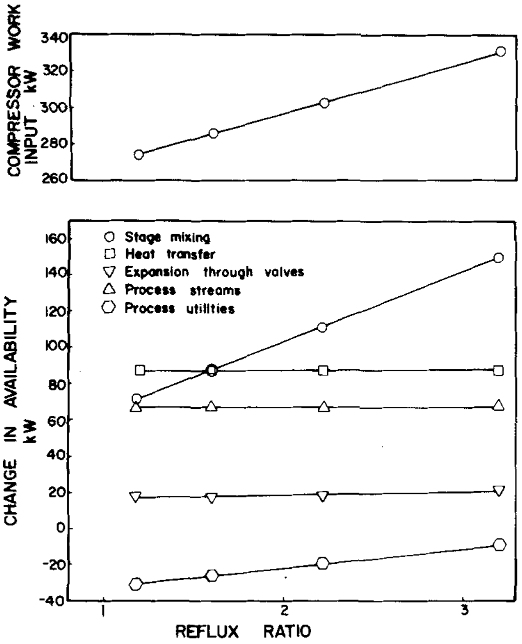

Figure 8. Availability changes in SRV distillation. Rectifier heat duty = 2.43 MW.

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NOTATION

B_i = $h_i - T_0S_i$, availability function of stream i, J/kg-mole
D = flow rate of the distillate, kg-mole/s
F = feed flow rate, kg-mole/s
h_b = molal enthalpy of the bottoms, J/kg-mole
h_d = molal enthalpy of the distillate, J/kg-mole
h_i = molal enthalpy of the material in stream i, J/kg-mole
h_L = molal enthalpy of the liquid, J/kg-mole
H_f = molal enthalpy of the feed, J/kg-mole
H_V = molal enthalpy of the vapor, J/kg-mole
L = flow rate of liquid leaving the rectifying section, kg-mole/s
m_i = flow rate of stream i, kg-mole/s
M_{syst} = the mass of the material contained within the system, kg-mole
q_j = rate of heat exchange between the system and an external heat reservoir maintained at temperature T_j, W
Q_R = cumulative rate of heat removal from the rectifying section including the condenser, W
Q_S = cumulative rate of heat addition to the stripping section including the reboiler, W
S_i = molal entropy of the material in stream i, J/kg-mole-K
S_{syst} = molal entropy of the material averaged over the system, J/kg-mole-K
t = time, s
T_B = reboiler temperature, K
T_C = temperature of the cold stream in a heat exchanger, K
T_D = condenser temperature, K
T_H = temperature of the hot stream in a heat exchanger, K
T_j = temperature of an auxiliary heat reservoir, K
T_o = temperature of the environment (infinite reservoir), K
T_R = temperature as a function of the height of the rectifying section, K
T_S = temperature as a function of the height of the stripping section, K
U_{syst} = molal internal energy of the material averaged over the system, J/kg-mole
V = flow rate of vapor entering the rectifying section, kg-mole/s
W = flow rate of the bottoms, kg-mole/s
w_l = work done per unit time by a part of the system bearing the label l, W
w_{min} = minimum work required for the (reversible) separation of a binary feed into a specified distillate and bottoms, J
x_d = mole fraction of the more volatile component in the distillate
x_f = mole fraction of the more volatile component in the liquid phase of the feed
y_f = mole fraction of the more volatile component in the vapor phase of the feed

Greek Letters

β_{hx} = availability loss arising from heat exchanges between streams, W
β_{sm} = availability loss arising from mixing of streams on distillation stages, W
β_v = availability loss arising from the passage of a material stream through a valve, W
η_i = thermodynamic efficiencies defined according to Equations (22)-(24)
σ = rate of irreversible entropy creation, W/K

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