

Energy optimization in multi-component distillation through unit modification

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In this study the available energy (exergy) method is used to analyze a usual distillation column from a thermodynamic point of view. The distillation unit consists of the column itself, the bottom reboiler and the top condenser. It is shown that, if an intermediate reboiler is introduced and its position specified, then the quantity of thermal energy to be supplied in order to achieve a designated separation, is reduced by approximately 10% which results due to optimizing the thermodynamic efficiency factor. The volume of the column is reduced by 13.5%, thus making this optimization more promising. The payback period of this modification is expected to be less than 12 months.

في هذا البحث تم الاستعانة بمبدأ الطاقة المتاحة والمتوفرة وتطبيقه في تحليل ودراسة وحدة تقطير للمركبات متعددة الأجزاء. وتتكون وحدة التقطير من عامود التقطير ومرجل إعادة الغلي في الأسفل والمكثف في الأعلى. ولقد تبين لنا، عند إدخال مرجل آخر للغلي في وسط عامود التقطير وتحديد مكان وجوده من الدرجة التي سيتم إدخاله عليها، أنه من الممكن أن تقل الطاقة اللازمة لتحقيق فصل المركبات متعددة الأجزاء بحوالي عشرة بالمائة، وهذا نتيجة تحسين معامل القدرة الحرارية. كما تم تقليل الحجم الكلي لوحدة التقطير بحوالي ثلاثة عشرة بالمائة، مما يعني أن عملية التعديل ستكون مجزية وواعدة. ومن المتوقع أن تكون مدة استرجاع الاستثمار لهذا التطوير لوحدة تقطير المركبات متعددة الأجزاء أقل من سنة واحدة.

Keywords: Multi-component distillation, Intermediate reboiler, Exergy, Thermodynamic efficiency optimization

1. Introduction

As the energy problem becomes more and more critical, the need for energy conservation wherever possible becomes more urgent. In the chemical processes industries (CPI), the major part of energy is consumed (used) in the separation systems [1]. Distillation is the most studied separation method. It is widely used in the petrochemical industry and in refineries, where the major part of energy is used in the distillation column. That is the reason why energy conservation trials are done in this section.

With the use of some thermodynamic concepts, especially of the exergy (available energy) a thermodynamic analysis of a distillation unit, consisting of the column, the bottom reboiler and the top condenser is carried out. This analysis indicates the positions, the reasons as well as the results of energy losses occurring in the distillation unit (process) of binary or multi-component mixtures. During the last years, the usual distillation unit configuration has been

modified into a configuration through which more efficient use of the supplied energy is achieved. Every configuration has as usual some advantages and disadvantages. If the new configuration results in the use of an intermediate reboiler, with the help of some thermodynamic criteria we can find:

- The most probable place of the intermediate reboiler introduction to the distillation column.
- The quantity of heat that should be delivered to the intermediate reboiler in order to achieve greater thermodynamic efficiency (less operational cost) and smaller column volume (less fixed cost)
- The reduction in consumed exergy and the percentage optimization of efficiency.

2. Energy conservation possibilities in distillation

The actual energy consumption in crude distillation is at least eight times the theoretical one [1]. It has been stated that 5% of the energy needed to achieve a desired

Table 1
Energy consumption in distillation

Product	Energy consumption kcal/kg	Product	Energy consumption kcal/kg
Cyclohexane	55	Ethylbenzene	1,268
Naphtha	74	Styrene	1,382
Formaldehyde	415	Ethylglycol	1,566
Methanol	658	Butadiene	1,765
Soda	1,050	Ethanol	5,008
Vinyl Chloride	1,226	Glycerin	8,330

distillation increases as the separation difficulty is increased (case of minimum difference of relative volatility or formation of azeotropes). Table 1 lists the energy consumption in distillation of some chemical products [2]. A distillation system is considered to be economically optimal when the overall operational and fixed costs have been reduced to a minimum [3,4].

2.1. Necessary background and discussion

Let Q denote any amount of heat and E any amount of energy added to or removed from the system. Applying the first law of thermodynamics (the law of energy conservation) yields:

$$Q_{in} = Q_{out} = Q_{feed} + Q_{reboiler}, \quad (1)$$

$$Q_{out} = Q_{bottom} + Q_{distillate} + Q_{condenser} + Q_{lost}, \quad (2)$$

and

$$E_{in} = E_{out}. \quad (3)$$

When the first law is applied, it is easily seen that 1kcal at 1000 °C and 1 kcal at 25 °C are the same. However, it is obvious that 1 kcal at 1000 °C is thermally more valuable than 1 kcal at 25 °C as it can be used for processing higher quality products from a thermodynamic point of view. So we need to apply the second law of thermodynamics (the law of energy degradation). A conversion factor is included that gives the fraction of energy transformed into useful work. Any transformation of energy into heat is considered as energy degradation [5].

2.2. Exergy

The maximum useful work Ex delivered by any system as it undergoes reversible

equilibrium with the surrounding environment is [3,6]

$$Ex = \Delta U + P_o \Delta V - T_o \Delta S - \sum_{i=1}^n \mu_{i,o} K_i. \quad (4)$$

Where k_i is the number of gram moles of the component i , U is the internal energy, P_o & T_o are the reference pressure and temperature respectively, V is the volume, S is the entropy, $\mu_{i,o}$ is the chemical potential of component i at reference conditions and n is the number of components in the system.

For steady-state conditions eq. (4) takes the following form:

$$Ex = \Delta H - T_o \Delta S - \sum_{i=1}^n \mu_{i,o} K_i. \quad (5)$$

For a thermally isolated system eq. (5) is further reduced into:

$$Ex = \Delta H - T_o \Delta S. \quad (6)$$

From eq. (6), one may conclude that the exergy of a system can be calculated using a Mollier diagram as the vertical distance of the straight line with slope T_o that passes through the points $\{P, T\}$ and $\{P_o, T_o\}$. The term $T_o \Delta S$ gives the part of energy that cannot be transformed into useful work. If we have a quantity of heat Q , then its exergy can be calculated as follows, since $dS = dQ/T$

$$Ex = \int (1 - T_o/T) dQ \quad (7)$$

And if T is constant, then

$$Ex = Q (1 - T_o/T). \quad (8)$$

For a fluid mixture of n components [7]

$$Ex' = RT_0 - \sum_{i=1}^n x_i \ln \gamma_i x_i \quad (9)$$

For a vapor mixture of n components

$$Ex' = RT_0 - \sum_{i=1}^n y_i \ln \phi_i y_i \quad (10)$$

For steady-state continuous operation, however,

$$\Delta Ex = \begin{matrix} \Delta H \\ 1 \rightarrow 2 \end{matrix} - T_0 \begin{matrix} \Delta S \\ 1 \rightarrow 2 \end{matrix} \quad (11)$$

In fig. 1, a reboiler is represented in which saturated steam of temperature T_s and latent heat of evaporation λ_v with flow rate V_v evaporates a saturated liquid at temperature $T_s - \Delta T_s$ and latent heat of evaporation λ_l and flow rate L_l .

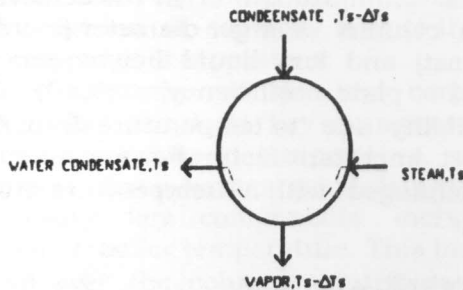


Fig. 1. A reboiler.

The exergy of the steam is

$$Ex_{(v)} = Q (1 - T_0 / T_s) = V_v \lambda_v (1 - T_0 / T_s) \quad (12)$$

And the exergy of the condensate is

$$Ex_{(l)} = Q [1 - T_0 / (T_s - \Delta T_s)] = V_v \lambda_l [1 - T_0 / (T_s - \Delta T_s)] \quad (13)$$

So the lost exergy is

$$Ex_{lost} = Ex_{(v)} - Ex_{(l)} = V_v \lambda_v T_0 [1 / (T_s - \Delta T_s) - 1 / T_s] \quad (14)$$

2.3. Reversibility and irreversibility

The irreversibility occurring in the distillation process are due to one or more of the following three reasons [2,8,6]:

- 1) Existence of temperature difference (ΔT not zero) in order to have heat exchanged between the various flows with different temperatures.
- 2) Existence of pressure difference (ΔP not zero), pressure drop due to flow along and within the distillation column.
- 3) Existence of concentration difference (Δc not zero) in order to have mass exchange between flows not in equilibrium.

The thermodynamic efficiency factor, J , indicates the reversibility or the degree of irreversibility of any process.

$$J = \text{Exergy Recovered} / \text{Exergy Consumed} = Ex_{(rec)} / Ex_{(cons)} = [Ex_{(cons)} - Ex_{(lost)}] / Ex_{(cons)} = 1 - \Delta Ex_{(lost)} / Ex_{(cons)} \quad (15)$$

If $\Delta Ex_{(lost)} = 0$ then $J = 1$ and the process is totally reversible.

2.4. Practical applications of exergy

In fig. 2 two diagrams are given, (A) is the Sankey exergy diagram and (B) is the Sankey enthalpy diagram, for crude oil distillation process [7].

2.4.1. Exergy analysis of distillation-minimum work of separation

The process of mixing increases the entropy of the system and reduces the available energy (exergy). So the separation of a mixture into its pure components requires the delivery of a quantity of exergy equal to the minimum work of separation. $Ex_{(rec)}$ in a distillation column is calculated as

$$Ex_{(rec)} = Ex_{(D)} + Ex_{(B)} + Ex_{(F)} \quad (16)$$

2.5. Required work of separation

In the distillation process the exergy is provided to the column at the reboiler at T_B and removed from the column at the condenser at T_D . It follows that:

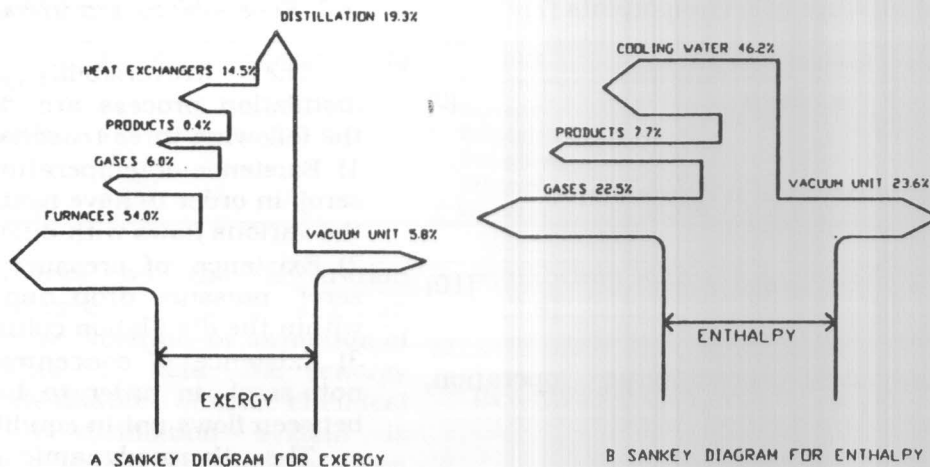


Fig. 2. A Sankey diagram.

$$Ex_{(reboiler)} = Q_B [(T_B - T_o) / T_B] , \quad (17)$$

and

$$Ex_{(condenser)} = Q_D [(T_D - T_o) / T_D] . \quad (18)$$

The difference between the above two eq. (17) and (18) gives the exergy consumed in the distillation column:

$$Ex_{(cons)} = Ex_{(reboiler)} - Ex_{(condenser)} = Q_B [(T_B - T_o) / T_B] - Q_D [(T_D - T_o) / T_D] . \quad (19)$$

In case when $Q_B = Q_D = Q$ (components with same amount of heat exchanged, then eq. (19) is transformed into:

$$Ex_{(cons)} = Q T_o [1 / T_D - 1 / T_B] . \quad (20)$$

The exergy consumed $Ex_{(cons)}$ is minimal when the distillation column operates with the minimum reflux ratio (R_{Dmin}). For any reflux ratio R_D other than R_{Dmin} we have:

$$Ex_{(cons)} = [Ex_{(cons)}]_{R_{Dmin}} [(1 + R_D) / (1 + R_{Dmin})] \quad (21)$$

If we consider that we have ΔT_D and ΔT_B temperature difference at the condenser and the reboiler, respectively, then

$$Ex_{(cons)} = Q_B [(T_B + \Delta T_B - T_o) / (T_B + \Delta T_B)] - Q_D [(T_D - \Delta T_D - T_o) / (T_D - \Delta T_D)] . \quad (22)$$

The thermodynamic efficiency factor J equals to $Ex_{(rec)} / Ex_{(cons)}$ and the exergy losses in distillation $\Delta Ex_{(lost)} = Ex_{(cons)} - Ex_{(rec)}$. The pressure drop, ΔP , is very effective in the reversibility phenomena and plays an important role when we have a large number of plates. Minimization of ΔP is achieved by using a column of larger diameter (increased fixed cost) and less liquid height per plate (reduced plate efficiency) [9,10]. The irreversibility due to temperature drop, ΔT , is the most important factor. For any amount of heat exchanged with ΔT temperature drop we have [7]:

$$\Delta Ex_{(lost)} = Q T_o \Delta T / T_{(av)}^2 , \quad (23)$$

where $T_{(av)}$ is the average temperature of the two temperatures with ΔT difference.

Fractional distillation is a dynamic reversible process, that can be carried out without exergy losses in case the reversibility conditions in all the plates all over the column can be achieved [11]. For a binary mixture this condition is satisfied when the operating lines coincide with the equilibrium curve on a McCabe-Thiele diagram [5,7,11,12]. In such a case we have equilibrium between the vapors and liquids at each plate all over the column, a condition requiring infinite number of intermediate reboilers and intermediate condensers as well as infinite number of plates as can be seen in fig. (3).

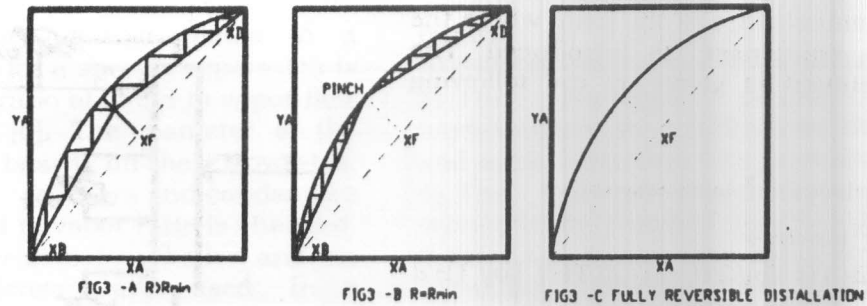


Fig. 3. McCabe-Thiele diagrams.

For a given separation between two key components of a multi-component mixture and a binary mixture consisting of the same two key components, the required work of separation of the multi-component mixture will be higher than that of the binary mixture, this is due to three main reasons [7,11]:

1) The non-key components of the multi-component mixture constitute part of the vapor and liquid streams. So, part of the thermal energy exchanged within the column will be used for evaporation, condensation and provision of sensible heat to these non-key components separation.

2) The existence of non-key components lighter than the light key component lowers the top condenser temperature, while the existence of non-key components heavier than the heavy key components increases the bottom reboiler temperature. This increase in ΔT all over the column contributes to more losses.

3) The existing concentration difference between the feed, vapor and liquid streams at the feed plate results in mixing where the entropy difference ΔS is increased [13,14].

3. Main results

3.1. Energy optimization methods in distillation

These are mainly divided into two groups [2,9,15]:

- 1) Methods of energy conservation with irreversibility minimization.
- 2) Methods of energy conservation with heat exchange irreversibility's minimization.

These two groups are further divided into sub-groups as:

- 1.a) use of intermediate reboilers and

condensers,

- 1.b) use of better filling materials and/or control,

and

- 2.a) introduction of vapor recompression
- 2.b) multiple effect distillation

Here, we focus on sub-group 1.a; i.e., the introduction of intermediate reboilers and condensers. We have seen that a method to minimize the irreversibility in a distillation column is to try to bring the vapor entering a plate more close to equilibrium with the liquid on that plate. Fig. 4 shows a diagram of a distillation column with an intermediate reboiler. It is assumed that, the heat load of the intermediate reboiler is $Q/2$ (half the load of the bottom reboiler).

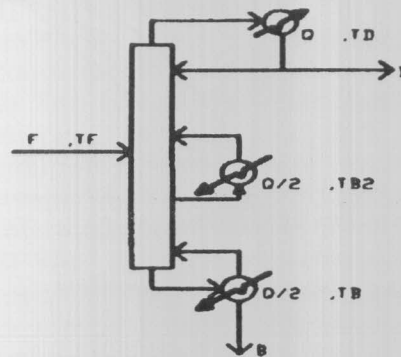


Fig. 4. Intermediate reboiler.

$$EX_{(reboiler)} = Q/2 [(T_B - T_0) / T_B], \quad (24)$$

$$EX_{(intermediate\ reboiler)} = Q/2 [(T_{B2} - T_0) / T_{B2}], \quad (25)$$

and

$$EX_{(condenser)} = Q [(T_C - T_0) / T_C], \quad (26)$$

T_{B2} is the temperature of the plate where the intermediate condenser is introduced. The exergy consumed is given by the following equation,

$$\begin{aligned}
 Ex_{(cons)} &= Ex_{(reboiler)} + Ex_{(intermediate\ reboiler)} \\
 &+ Ex_{(condenser)} \\
 &= Q/2((T_B T_o)/T_B) + Q/2((T_{B2} T_o)/T_{B2}) \\
 &+ Q((T_C T_o)/T_C) \\
 &= Q/2((1 T_o/T_B) + (1 T_o/T_{B2})) \\
 &+ Q((1 T_o/T_C)) \\
 &= T_o(Q/TC + 0.5Q/T_B + 0.5Q/T_{B2}). \quad (27)
 \end{aligned}$$

Then the minimization in energy consumption when using an intermediate reboiler is,

$$\Delta Ex_{(cons)} = Q/2 T_o (1 / T_{B2} - 1 / T_B). \quad (28)$$

Figs. 5-a, and 5-b show the McCabe-Thiele diagrams for a distillation column with intermediate reboiler operating at $R = R_{min}$ and $R > R_{min}$, respectively. While fig. 6 shows the diagram of a distillation column with intermediate condenser [10,12]. Suppose that the heat load of the intermediate condenser is

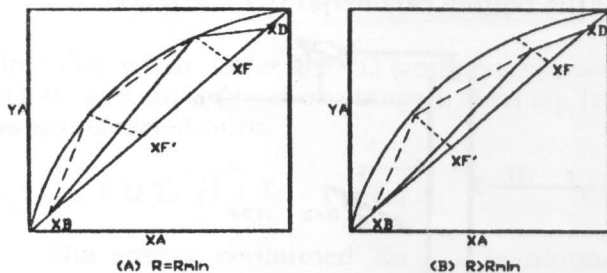


Fig. 5. McCabe- Thiele diagrams with intermediate reboiler.

$Q_{D2} = Q/2$, if T_{D2} is the temperature of the plate where the intermediate condenser is introduced, then the exergy consumed given by eq. (26) reduces to,

$$\begin{aligned}
 Ex_{(cons)} &= Q_B (1 - T_o / T_B) - Q_{D1} (1 - T_o / T_{D1}) \\
 &- Q_{D2} (1 - T_o / T_{D2}) \\
 &= T_o [Q_{D1}/ T_{D1} + Q_{D2} / T_{D2} - Q_B/T_B]. \quad (29)
 \end{aligned}$$

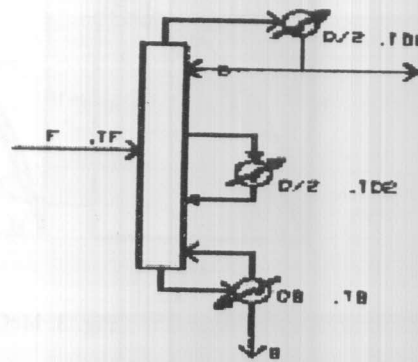


Fig. 6. Intermediate condenser.

and the minimization in energy consumption when an intermediate condenser is introduced is,

$$\Delta Ex_{(cons)} = Q/2 T_o (1 / T_{D1} - 1 / T_{D2}) \quad (30)$$

Figs. 7-a and 7-b show the McCabe-Thiele diagram for a distillation column with intermediate condenser operating at $R = R_{min}$ and $R > R_{min}$, respectively [10].

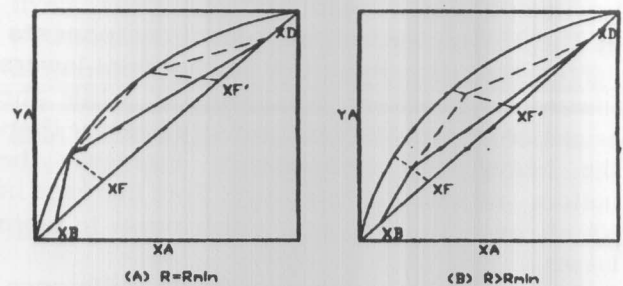


Fig. 7. McCabe- Thiele diagrams with intermediate condenser.

Observe that for a specific value of heating load in the condenser, where the distribution of the components and the flow rates of the feed and products are constant, the overall heat given in the stripping section has one unique value irrelevant to the number of reboilers used [8, 9]. In this case, the thermodynamic efficiency is increased (exergy consumption is minimized). The same is true in case an intermediate condenser is introduced. In this case the overall heat to be removed in the enrichment section has one unique value irrelevant to the number of intermediate condensers used.

The number of required plates in a distillation column for a specific separation is determined by the ratio of liquid to vapor flow inside the column [6]. The diameter of the column chosen is based on these flowrates. When intermediate reboilers and condensers are used, the liquid to vapor ratio is changed. So when an intermediate reboiler and an intermediate condenser are used in a distillation column, we need to reduce the diameter of the column above the intermediate condenser and below the intermediate reboiler as a result of reduced liquid and vapor flow ratios in these two parts of the column. Consequently, the distillation column can look like the one shown in fig. (8).

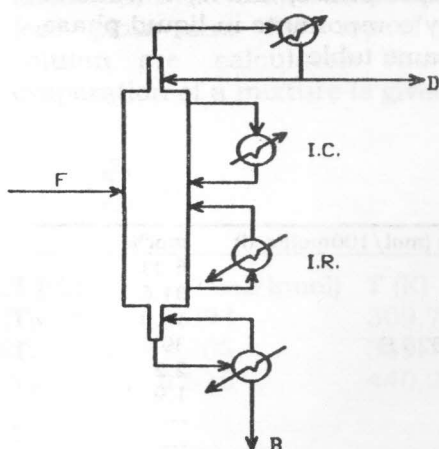


Fig. 8. Distillation column with intermediate reboiler and condenser.

3.2. Cases when intermediate reboilers or condensers are indicated [2, 4, 6, 8, 16]:

1) The use of an intermediate reboiler is indicated when there exists available thermal energy (steam at various pressures) at temperatures suitable to be used in the intermediate part of the column.

2) The use of intermediate condenser is more beneficial than the use of intermediate reboiler as cooling costs higher than heating. In cryogenic distillation systems the use of intermediate condenser is indicated because it is not expensive as cooling at the top condenser. In the case of distillation at temperatures above that of the environment, the use of intermediate condensers is economically justified only when the heat

removed by the intermediate condenser can be further used in the process [16].

3) The pay-back period for these modifications is usually less than one year and sometimes even less than six months.

4) The use of intermediate reboilers is economically justified for $T_B > 150$ °C and x_{lk} (light key component concentration is less than 0.3. On the other hand, the use of intermediate condensers is economically justified for $T_B < 40$ °C and x_{hk} (heavy key component concentration) is greater than 0.3.

5) The increased load of intermediate systems change the height, diameter as well as the heat exchange area of the column [8].

6) The use of less expensive thermal energy levels lowers the operational and energy costs resulting in a shift of the minimum operational cost into higher reflux ratios.

Generally, when we need to find the suitable plate for the introduction of any intermediate reboiler or condenser; (a suitable place is considered to be the one that produces optimal combination of energy consumption, column volume and cost of intermediate condenser and/or intermediate reboiler), and the temperature at which the heat load is to be provided; the following remarks should be taken into consideration:

1) More theoretical plates in the column are required when an intermediate reboiler or condenser is introduced.

2) The intermediate reboiler (with the same energy consumption and volume of the column) associated with the lowest temperature is preferred since the intermediate reboiler is of minimum cost.

3) The intermediate reboiler (when a heating medium of sufficient temperature is available) is provided with a specific quantity of heat in order to achieve a better combination of energy consumed and column volume.

4. Example from bibliography

The idea of using an intermediate reboiler was applied to an example given by Robinson and Gilliland [17]. It is an example of a gasoline stabilization unit, the feed composition of which is given in table 2. It is desired to recover 96% of the normal butane (n-butane) in the bottom residue while the

same residue should not contain more than 0.25% of the propane fed to the column. The plate where the intermediate reboiler is introduced should be determined together with the concept of consumed energy minimization. A reflux ratio of 2 is to be maintained in the upper part of the column and the feed should be introduced at the plate below of which $L/V = 1.5$. Observe that the concentrations of the lighter than propane in the residue as well as the concentration of the heavier than the n-butane in the distillate are negligible. Also, 20% of the iso-butane [which is lighter than the n-butane] is in the distillate. The light key component (lk) is propane, and the heavy key component (hk) is n-butane.

Because the distillate is highly volatile, it is withdrawn in the vapor phase, and a partial condenser is used to provide the required

reflux. The modified method of Lewis-Matheson is used in which the term y_i is given by the following equation:

$$y_i = (a_{ih} X_i) / \left(\sum_{i=1}^n a_{ih} X_i \right), \quad (31)$$

where a_{ih} is the relative volatility of component i with respect to that of the heavy key component hk .

Tables 3-a and 3-b give the distribution of the various components fed into the column as given by Robinson and Gilliland. Also T_j , $k_n C_4$, y_l/y_h (ratio of light to heavy key components in vapor phase) and x_l/x_h (ratio of light to heavy key components in liquid phase) are given in the same table.

Table 2
Composition of feed, residue and distillate for the process

Feed	mol%	Residue (mol/100molFeed)	mol%	Distillate (mol/100molFeed)	mol%
CH ₄	2.0	---	---	2.0	6.33
C ₂ H ₆	10.0	---	---	10.0	31.6
C ₃ H ₈	6.0	---	---	6.0	19.0
lk C ₃ H ₈	12.5	0.0025 B	0.25	12.5-0.0025 B	39.0
i-C ₄ H ₁₀	3.5	2.8	4.10	0.7	2.2
hk n-C ₄ H ₁₀	15.0	14.4	21.1	0.6	1.9
C ₅	15.2	15.2	22.2	---	---
C ₆	11.3	11.3	16.5	---	---
C ₇	9.0	9.0	13.2	---	---
C ₈	8.5	8.5	12.4	---	---
> C ₈	7.0	7.0	10.2	---	---
		68.2+0.0025B		31.8-0.0025B	
		68.2 + 0.0025B = B			
F=100 mol	B=68.4 mol	D= 31.6 mol	P=1 atm	R _{min} =0.75	x_l/x_h =0.833

Table 3-a
Component distribution in plate j (vapor phase)

y_i	y_D	y_1	y_2	y_3	y_4	y_5	y_6	y_7	y_8	y_9	y_{10}	y_{11}	y_B
C ₁	.0633	0.025	.022	.022	.022	---	---	---	---	---	---	---	---
C ₂	0.316	0.204	0.157	0.14	0.133	0.05	---	---	---	---	---	---	---
C ₃	0.19	0.21	0.195	0.169	0.141	0.12	---	---	---	---	---	---	---
lk C ₃	0.39	0.463	0.452	0.4	0.338	0.321	0.278	0.19	0.122	.0735	.041	.022	.01
iC ₄	.022	.0463	.074	.0997	0.104	0.131	0.179	0.191	0.193	0.184	0.161	0.134	.0955
hk nC ₄	0.19	.0503	0.100	0.169	0.208	0.295	0.425	0.487	0.535	0.56	0.552	0.518	0.42
C ₅	---	---	---	.015	.042	.056	.0795	.0905	0.106	0.129	0.165	0.25	0.257
C ₆	---	---	---	.0017	.011	.016	.0235	.0255	.0285	.035	.0535	.0775	0.125
C ₇	---	---	---	---	.004	.0065	.0095	.0105	.011	.0125	.0175	.026	.0575
C ₈	---	---	---	---	.002	.003	.0045	.0045	.005	.0055	.008	.011	.03
> C ₈	---	---	---	---	.0003	.0005	.0005	.001	.001	.001	.002	.0025	.008
KnC ₄	0.288	0.356	0.41	0.56	0.76	0.833	1.075	1.14	1.2	1.28	1.33	1.48	1.99
T _j °C	36.7	48.9	54.4	72.8	86.7	96.1	116.1	121.1	124.4	129.4	133.9	140.6	167.2
y_l/y_h	20.5	9.2	4.49	2.36	1.625	1.088	0.654	0.39	0.228	0.131	.0743	.0425	.0238

Table 3-b
Component distribution in plate j (liquid phase)

X _i	X _D	X ₁	X ₂	X ₃	X ₄	X ₅	X ₆	X ₇	X ₈	X ₉	X ₁₀	X ₁₁	X _B
C ₁	.006	.002	.0014	.0014	.0014	---	---	---	---	---	---	---	---
C ₂	0.148	.0755	.0515	.042	.0333	.012	---	---	---	---	---	---	---
C ₃	0.22	0.197	0.158	0.117	.08	.061	---	---	---	---	---	---	---
lk C ₃₊	0.5	0.483	0.405	0.312	0.215	0.186	0.127	.0822	.0498	.0282	.0155	.0075	.0025
iC ₄	.0585	0.1	0.138	0.145	0.101	0.133	0.141	0.142	0.136	0.121	0.103	.0773	.041
hk nC ₄	.066	0.142	0.245	0.303	0.267	0.354	0.395	0.417	0.424	0.438	0.415	0.351	0.211
C ₅	---	---	.0225	.0625	0.111	0.127	0.134	0.145	0.16	0.184	0.214	0.245	0.222
C ₆	---	---	.0025	.0165	.0657	0.707	.072	.074	.0783	.0907	0.107	.138	0.165
C ₇	---	---	---	.006	.0483	.0503	.051	.0513	.0523	.0557	.0613	.0823	0.132
C ₈	---	---	---	.003	.0437	.0447	.0447	.045	.0453	.047	.049	.0617	0.125
> C ₈	---	---	---	.0004	.0343	.0345	.0347	.0348	.0349	.0353	.0357	.0393	0.102
X _i /X _h	7.576	3.41	1.656	1.03	0.804	0.526	0.322	0.193	0.112	.0644	.0373	.0214	.0118

4.1. Exergy consumption in the original column

Based on the data in table 4 [18], the latent heat of evaporation at the bottom, the feed plate and the top of the distillation column are calculated. The latent heat of evaporation of a mixture is given by,

$$\lambda_{\min} = \sum_{i=1}^n x_i \lambda_i \quad (32)$$

T (°C)	λ (kcal/kmol)	T (K)
T _D 36.7	2,393	309.7
T ₄ 86.7	3,205	359.7 (T _o =25 °C=298 K)
T _B 167.2	4,515	440.2

Table 4
Latent heat of evaporation (kcal/kmol) for each component

	36.7 °C	167.2 °C	102.2 °C
C ₁	616	112	364
C ₂	1,961	1,232	1,597
C ₃	2,633	1,961	2,297
lk C ₃₊	2,913	2,241	2,577
iC ₄	3,249	2,745	2,997
hk nC ₄	3,473	2,969	3,221
C ₅	4,089	3,697	3,893
C ₆	4,762	4,313	4,537
C ₇	5,662	5,098	5,350
C ₈	6,666	5,994	6,330
> C ₈	8,347	7,562	7,955

The quantity of heat to be provided at the column's bottom reboiler is,

$$Q_B = V \lambda_B = 136.8 \times 4,515 = 617,500 \text{ kcal/100 kmol Feed}$$

The quantity of heat to be removed at the column's top condenser is:

$$Q_D = L_o \lambda_D = 63.2 \times 2,393 = 151,240 \text{ kcal/100 kmol Feed}$$

The exergy consumption in the column is,

$$EX_{(cons)} = Q_B (1 - T_o/T_B) - Q_D (1 - T_o/T_D) = 193,800 \text{ kcal/100 kmol Feed}$$

4.2. Exergy consumption in the column with intermediate reboiler at 120 °C

The ratio L₁/V₁ and the reflux ratio in part 1 of the column shown in fig. 10 are:

$$L_1/V_1 = 159.6/91.2 = 1.75$$

$$L_1/D = 159.6/31.6 = 5.05$$

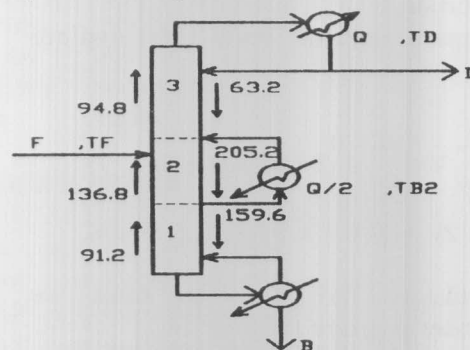


Fig. 10. Vapor and liquid flows in column with intermediate reboiler in kmol.

The quantity of heat provided by the bottom's reboiler

$$Q_{B1} = 2Q_B/3 = 411,680 \text{ kcal/100 kmol Feed.}$$

The quantity of heat provided by the intermediate reboiler at 120 °C

$$Q_{B2} = Q_B/3 = 205,840 \text{ kcal/100 kmol Feed.}$$

The quantity of heat removed by the top's condenser is the same as that calculated for the original column:

$$Q_D = L_o \lambda_D = 63.2 \times 2,393 = 151,240 \text{ kcal/100 kmol Feed.}$$

The exergy consumption in the column with intermediate reboiler is,

$$\begin{aligned} EX_{(cons)} &= Q_{B1} (1 - T_o/T_{B1}) + Q_{B2} (1 - T_o/T_{B2}) \\ &\quad - Q_D (1 - T_o/T_D) \\ &= 132,990 + 49,760 - 5,710 \\ &= 177,040 \text{ kcal/100 kmol Feed.} \end{aligned}$$

This consumption is lower by 16,760 kcal/100 kmol Feed than that in the original column.

4.3 Calculation of the column's volume with and without an intermediate reboiler

Let the column's volume without the intermediate reboiler be V, then

$$V = \left(\sum_1^3 V + \sum_4^{11} V \right) K = [(94.8 \times 3) + (93.6 \times 8)] K = 1.379 K,$$

where K is a constant.

The column's volume with intermediate reboiler Vr' is,

$$\begin{aligned} Vr' &= \left(\sum_1^3 V + \sum_4^5 V + \sum_6^{12} V \right) K = [(94.8 \times 3) \\ &\quad + (136.8 \times 2) + (91.7 \times 7)] K = 1,196 K. \end{aligned}$$

Then the volume reduction in case we introduce an intermediate reboiler is,

$$(1,379 - 1,196) / 1,379 = 13.3\%.$$

4.4 Calculation of recovered exergy in the distillation column

For purposes of simplification, it is assumed that the feed, residue and distillate

are ideal mixtures; so their activity coefficients are equal to unity.

The exergy of an ideal mixture based on its composition is given by [7],

$$EX = R T_o \sum x_i \ln x_i.$$

The recovered exergy by the distillation column is given by

$$EX_{(rec)} = EX_B + EX_D - EX_F,$$

where B, D and F are the bottom residue, top distillate and feed, respectively.

$$\begin{aligned} EX_{(rec)} &= R T_o \{ (B \sum x_{iB} \ln x_{iB}) \\ &\quad + (D \sum x_{iD} \ln x_{iD}) - (F \sum x_{iF} \ln x_{iF}) \} \\ &= 1.987 \times 298 \times 75.16 \\ &= 33,850 \text{ kcal/100 kmol Feed.} \end{aligned}$$

5. Conclusions and suggestions

This paper introduces the exergy calculations for a usual distillation column and that of the same column with an intermediate reboiler. Also, the calculations for the consumed energy and the volume are given. It is shown in table 5 that the introduction of an intermediate reboiler into the existing distillation column has produced some noticeable results:

- 1) The thermodynamic efficiency factor J was optimized by 10%. This has as a consequence a substantial reduction in operational cost
- 2) The volume of the distillation column was reduced by 13.5% although the number of plates was increased by one. This produces a reduction in a factor affecting the fixed cost of the column.
- 3) The reflux ratio as well as the recovered exergy remained the same but the consumed exergy was reduced by 16,760 kcal/100 kmol Feed.

It is suggested to investigate the possibility of applying a heat pump in the distillation column and to study also a combination of a heat pump and an intermediate reboiler and/or intermediate condenser. In a final phase the cost of such a modification should be analyzed and the payback period of such a revamp determined.

Table 5
Table of results

	Ordinary column	Column with intermediate reboiler
Plates above feed	3	3
Plates below feed	9	10
Feed plate	4	4
Plate of intermediate reboiler	---	5
Reflux ratio L/D	2.0	2.0
Reflux ratio L ⁻ /D	6.49	5.05
L ⁻ /V ⁻ (stripping section)	1.5	1.5 & 1.75
Column's volume reduction	---	13.3 %
Q Bottom Reboiler	617,520 kcal	411,680 kcal
Q Intermediate Reboiler	---	205,840 kcal
Q Condenser	151,240 kcal	151,240 kcal
Overall heat supply	617,520 kcal	617,520 kcal
EX Bottom Reboiler	199,500 kcal	132,990 kcal
EX Intermediate Reboiler	---	49,760 kcal
EX Condenser	5,710 kcal	5,710 kcal
Exergy _(rec)	33,850 kcal	33,850 kcal
Exergy _(cons)	193,790 kcal	177,040 kcal
Reduction in exergy _(cons)	---	16,750 kcal
Thermodynamic efficiency factor, J	17.5%	19.1%
J optimization, %	---	9.1%

Nomenclature

a	Relative volatility
B	Bottom product
D	Distillate
E	Energy
Ex	Exergy
Ex'	Specific exergy
F	Feed
J	Thermodynamic efficiency factor
H	Enthalpy
K	Constant for volume calculations
k _i	Number of grammoles of component i
L _L	Liquid flow rate
n	Number of components in the mixture
P	Pressure
P _o	Reference pressure
Q	Heat
R _{min}	Minimum reflux ratio
R	Universal gas constant
S	Entropy
T	Temperature
T _o	Reference temperature
T _s	Steam temperature
U	Internal energy
V	Volume
V _v	Vapor flow rate
V _{R'}	Volume with inter. reboiler
x _i	Mole fraction of i in liquid phase
y _i	Mole fraction of i in vapor phase

Subscripts

(av)	average or mean value
B	bottom residue
C	condensate
(cons)	consumed
D	top distillate
F	feed
ih	i relative to h
in	inlet
hk	heavy key
l	liquid
lk	light key
(lost)	lost
min	minimum
out	outlet
(rec)	recovered
v	vapor

Greek letters

γ _i	activity coefficient
Δc	concentration difference
ΔP	pressure difference
ΔT	temperature difference
μ _{i,o}	chemical potential of i
λ	latent heat of evaporation
Σ	sum of
φ _i	fug. coefficient of i in vapor phase

Prefix

Δ change or difference

Suffix

~ modified

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