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Energy saving and thermodynamic efficiency of a double-effect distillation column using internal heat integration

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Abstract−The existing internally heat-integrated distillation column with the problem of utilizing a compressor is modified to propose a new heat-integrated distillation column without the compressor. Two identical columns of a conventional binary distillation are implemented to the heat integration. The energy used in the reboiler is recovered by the internal heat integration between the stripping section of one of the columns at lower pressure and the rectifying section of the other higher pressure column. The heat integration is similar to double-effect distillation, but internal heat integration requires less pressure elevation. The performance of energy saving and thermal efficiency improvement of the proposed system is evaluated with the two examples of the benzene-toluene and methanol-ethanol processes. The performance comparison indicates that the proposed system requires 17.4% less of reboiler duty for the benzene-toluene process and 15.8% less of heating duty for the methanol-ethanol process. The thermal efficiencies are 16.3% and 23.8% for the benzene-toluene and methanol-ethanol processes, respectively. Elimination of the compressor makes the column operation easy and the separate reboilers and condensers for the two columns in the proposed system provide flexible control, when the controllability of the proposed system is compared with that of the existing internally heat-integrated distillation column.

Key words: Heat-integrated Distillation, Energy-efficient Distillation, Internal Heat Integration, Double-effect Distillation

INTRODUCTION

In a conventional distillation column there are two heat exchangers—a reboiler and a condenser. Though one of them consumes heat and the other releases heat, it cannot be recycled due to the negative temperature difference between them. In practice the heat integration occurs between different distillation columns after their operating temperatures are consulted. A heat-integrated distillation column within the column, the internally heat-integrated distillation column (HIDiC) [1,2], utilizes the energy removed from the upper section of the binary column, rectifying section, in order to heat its lower section, stripping section. This heat integration is accomplished by using the two sections in a single column sharing the same column wall for heat transfer. One problem in the scheme is that the temperature in the upper section is commonly lower than in the lower section, while the heat transfers from the upper to the lower. Therefore, vapor compression is necessary for the temperature elevation of the upper section, and the utilization of a compressor involves difficulties in its cost, operation and maintenance [3]. An ideal HIDiC does not require external heat supply except for the electricity used by the compressor, but the heat exchangers are necessary for the startup and control of the column. Recently, the heat-integration with external heat exchangers has been proposed: between condenser and reboiler [4] and between trays in rectifying and stripping sections [5,6]. These systems use two or three heat exchangers, while the HIDiC has the heat exchanger in every paired tray.

When two distillation columns process a half of the feed each in a double-effect distillation column [7], the heat integration between the reboiler of the first column and the condenser of the second is available by increasing the operating pressure of the second column. Unlike the HIDiC the heat integration is implemented in the half of the distillation system, but no compressor is necessary. One problem in the heat integration is the large temperature difference between the condenser and reboiler, leading to the large elevation of the operating pressure of the second column. Note that the condenser temperature is the lowest in a conventional column and the reboiler is the highest. The tray-by-tray heat integration as in the HIDiC can be employed here for the reduction of the large pressure increase [8,9]. The tray-by-tray heat integration indicates the utilization of a pair of diabatic distillation columns. In a recent study a structured column was introduced to increase the heat transfer rate in the heat-integrated columns [10]. The application of the HIDiC has also been extended to batch distillation [11] and reactive distillation [12,13]. The diabatic distillation also reduces the exergy loss from the separation process [14,15]. Exergy is the maximum amount of work obtainable. A reversible process gives the maximum in a state of thermodynamic equilibrium. The reduction of the exergy loss has been pursued for the improvement of thermodynamic efficiency in distillation columns by utilizing two columns [16], intermediate heat exchanger [17] and diabatic column section [18].

Though the energy conservation of a divided wall column (DWC) is well known to be implemented in many industrial processes [19-21], its application is limited to the ternary distillation systems. In other words, two conventional distillation columns in sequence are replaced with the DWC. In practice, the replacement of a currently operated two-column system with a totally new column is not welcomed by field operators due to the lack of experience with the new system. The introduction of a new binary system requiring less energy demand is much easier than that of the DWC for ternary separation.

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In this study, a double-effect distillation system utilizing an internally heat-integrated column is proposed to replace the conventional binary column for the reduction of energy requirement, and its design and operation are explained. The performance of energy conservation of the proposed system is examined with two example processes. In addition, the thermal efficiency of the proposed system is compared with that of the conventional distillation system.

COLUMN STRUCTURE AND DESIGN

In a binary distillation column the heat integration is accomplished by operating columns at different pressures [1,2,22] as shown in Fig. 1, and it is known as the internally heat-integrated distillation column (HIDiC). In the practical application the difficulty caused by the compressor utilization has been indicated from the field engineers to introduce a new HIDiC without the compressor utilization [3]. Instead of raising the operating pressure of the rectifying section, the operating pressure of the stripping section is reduced by using a vacuum pump between the two sections of the heat integration.

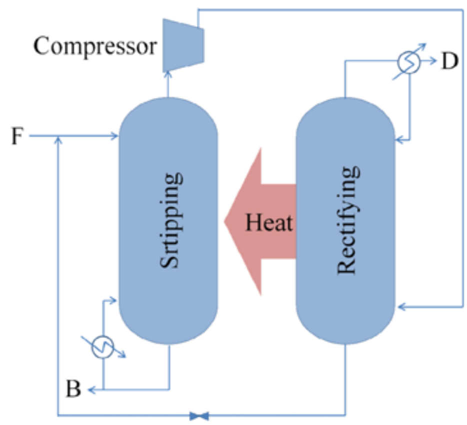


Fig. 1. Schematic diagram of an internally heat-integrated distillation column.

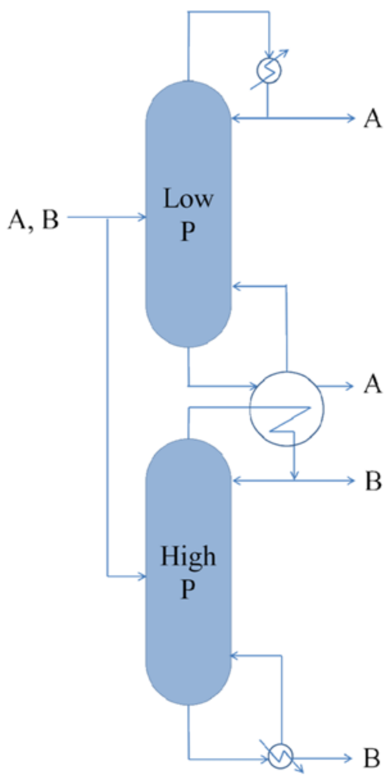


Fig. 2. Schematic diagram of a double-effect distillation column.

A direct heat integration [7] is possible between the reboiler of the first column and the condenser of the second column of two identical binary columns operated at different pressures as shown in Fig. 2. The second column is operated at the higher pressure, and the reboiler of the first column and the condenser of the second are combined into a single heat exchanger. A similar distillation structure of double-effect distillation was implemented in a two-column system of a binary separation [7] and a ternary separation [23]. Because the temperature difference of two heat exchangers is large, the pressure elevation of the second column is significant, resulting in the large reduction of thermodynamic and tray efficiencies of the second distillation column. The tray-by-tray heat integration as demonstrated in Fig. 3 lowers the pressure difference between the two columns, and the diabatic operation, though a half of each column involved, improves the efficiencies of the distillation system. The improvement will be calculated later. The tray-by-tray heat integration requires less temperature increase in the second column compared with the direct heat integration of Fig. 2. The conceptual heat integration of Fig. 3 can be materialized by the introduction of an internally heat-integrated distillation column as given in Fig. 4. Note that the heat integration occurs partially. The rectifying section of the first distillation column and the stripping section of the second are left as they are.

The structural design of the proposed distillation system is directly adopted from the original column except the internally heat-integrated column. Namely, the first and last sections of the system

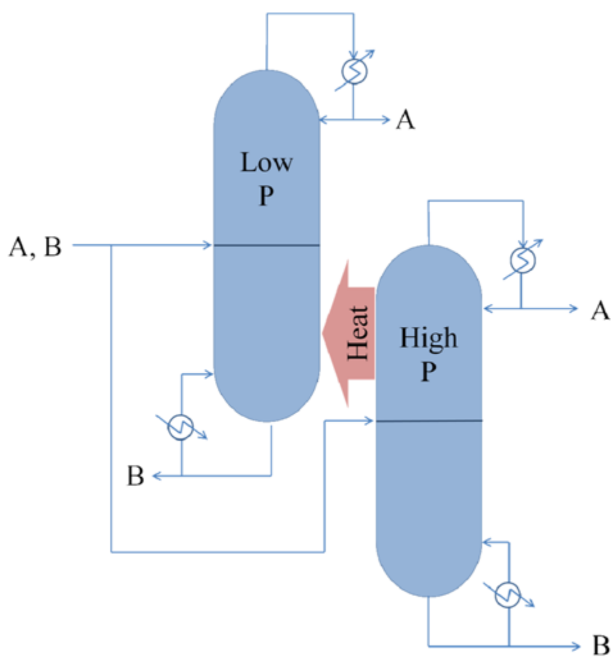


Fig. 3. A conceptual schematic of a double-effect internally heat-integrated distillation column.

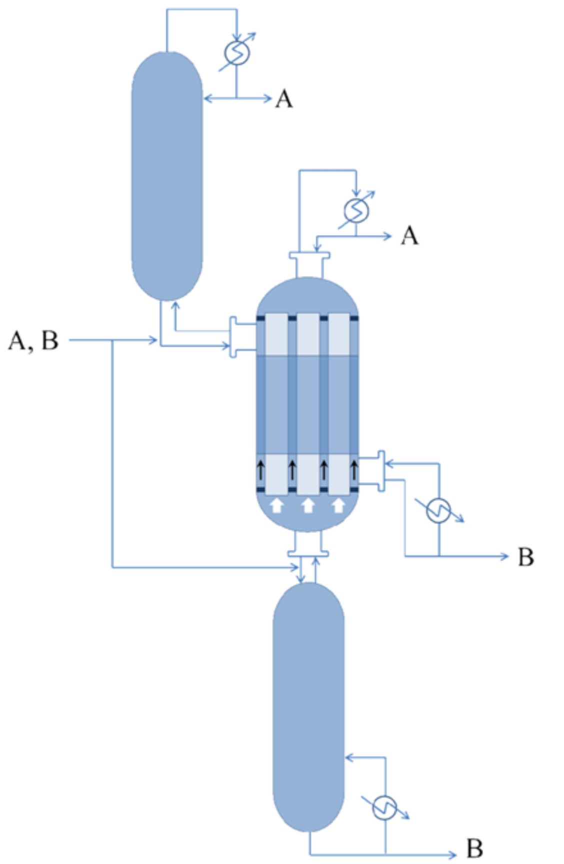


Fig. 4. Schematic diagram of a double-effect internally heat-integrated distillation column for practical application.

use those of the sections in the numbers of trays of the original column. In practice, the existing column can be utilized as one of the sections, though the number of trays is larger than the design of the new system. Unless the numbers of trays in the rectifying and stripping sections of the original distillation column are quite different, the number of trays of the heat integrated column can be determined as the smaller number between the two sections.

The optimum design of the proposed system is found from the maximum recovery of the removed heat from the rectifying section of the second column. Therefore, the split of feed can be a half to each stream. In the design of the proposed system, the commercial design software HYSYS was used, and a series of heat exchangers were installed for the heat transfer between the pairing trays in the heat-integrated column. The overall heat transfer coefficient and heat transfer area are given as 1 kW/m^2 °C and 150 m^2 per tray, respectively, after consulting the design guideline in Gadalla et al. [24]. The effect of the heat transfer coefficient UA on the heat transfer is nearly constant, when its value is larger than 2.5 kW/°C for the debutanizer at a feed rate of 227 kmol/h [25,26]. Because the value of UA in this study is larger than that, the UA value does not affect the overall performance in the proposed system and double-effect distillation column. The installation of the additional panels for heat transfer in a concentric column significantly increases the heat transfer area. For the column with a throughput of 100 kmol h^{-1}, the heat transfer area is available up to 621 m^2 per tray in the design of the previous study.

EXERGY LOSS AND THERMAL EFFICIENCY

The maximum available work from a reversible process is represented as exergy, and it is calculated from the enthalpy and entropy of the process.

$$E=(H-H_o)-T_0(S-S_o) \tag{1}$$

where the subscript o indicates the standard state of 25 °C and 1 atmospheric pressure used here. In a tray of a distillation column the exergy balance is given as

$$\dot{E}_{loss}=L_{j-1}E^L_{j-1}+V_{j+1}E^V_{j+1}+F_jE^F_j-L_jE^L_j-V_jE^V_j-S_jE^S_j-\dot{E}_{Q,j} \tag{2}$$

where the thermal exergy of energy stream is found from

$$\dot{E}_{Q,j}=Q_j\left(1-\frac{T_o}{T}\right) \tag{3}$$

The exergy balance (2) can be used with slight modification for the condenser and reboiler of the distillation column. The left hand side of the balance is the rate of exergy loss, which indicates the irreversibility of the process. Because the tray-by-tray computation of exergy transfer is complicated, a diagram of the Carnot factor and enthalpy is useful to calculate the exergy loss in a distillation column [27]. The Carnot factor is defined as

$$f=\left(1-\frac{T_o}{T}\right) \tag{4}$$

The thermodynamic efficiency of the column is defined as [15, 28]

$$\eta=\frac{\dot{W}_m}{\dot{W}_m+\dot{E}_{loss}} \tag{5}$$

where the minimum work is calculated from [29]

$$\dot{W}_m=DE^D+BE^B-FE^F \tag{6}$$

Though the thermodynamic efficiency of a diabatic distillation column utilizing in-tray heat transfer like the tray-by-tray heat transfer of this study is higher than that of adiabatic, the energy requirement of the diabatic column is not significantly lower than that of the adiabatic column [30].

EXAMPLE PROCESSES

Two example processes are used to examine the performance of energy saving and exergy loss of the proposed distillation system over the conventional column. The separation of benzene-toluene mixture was used for the study of the HIDiC [2] and was applied in this study. The mixture of methanol and ethanol was used as the second example system. The Peng-Robinson EOS was implemented for the calculation of VLE of the first system, and the NRTL equation was used for the second. The structural and operating information of the proposed and conventional systems for the benzene-toluene mixture is listed in Table 1. That of the methanol-ethanol mixture is summarized in Table 2.

RESULTS AND DISCUSSION

The heat removed from the rectifying section of a distillation col-

Table 1. Structural information, operating conditions and compositions in the proposed, existing double effect and conventional distillation systems for the benzene-toluene process. Tray numbers are counted from the top

Name	Proposed column		Double effect		Conventional
	1st	2nd	1st	2nd	
Structural					
Number of trays	20	20	20	20	40
Feed tray	10	10	10	10	20
Number of heat-integrated trays	10	10			
Operating					
Pressure (kg/cm^2)-top	1.0	2.0	1.0	3.0	1.0
Temperature (°C)					
Top	81.4	105.9	80.3	125.4	81.4
Feed tray	93.6	115.1	91.3	138.9	94.2
Bottom	109.6	133.3	108.0	147.7	111.8
Feed (kmol/h)	400	400	400	400	800
Overhead (kmol/h)	200	200	200	200	400
Bottom (kmol/h)	200	200	200	200	400
Reflux (kmol/h)	300.5	28.0	328.4	-	526.0
Vapor boilup (kmol/h)	82.5	627.0	-	319.2	880.5
Cooling duty (GJ/h)	15.4	6.76	16.2	-	28.6
Reboiler duty (GJ/h)	2.74	19.9	-	9.9	29.1
Preheater (GJ/h)		1.4		15.0	
Composition (mol frac.)					
Feed	0.5/0.5	0.5/0.5	0.5/0.5	0.5/0.5	0.5/0.5
Product-benzene	0.950	0.950	0.975	0.925	0.950
-toluene	0.951	0.946	0.975	0.925	0.950

Table 2. Structural information, operating conditions and compositions in the proposed, existing double effect and conventional distillation systems for the methanol-ethanol process. Tray numbers are counted from the top

Name	Proposed column		Double effect		Conventional
	1st	2nd	1st	2nd	
Structural					
Number of trays	24	24	24	24	40
Feed tray	12	12	12	12	20
Number of heat-integrated trays	12	12			
Operating					
Pressure (kg/cm^2)-top	1.0	1.8	1.0	3.0	1.0
Temperature (°C)					
Overhead	64.7	80.3	64.2	95.6	64.7
Feed tray	72.6	85.9	70.6	102.0	72.4
Bottom	78.8	93.4	76.9	106.5	80.9
Feed (kmol/h)	400	400	400	400	800
Overhead (kmol/h)	200	200	200	200	400
Bottom (kmol/h)	200	200	200	200	400
Reflux (kmol/h)	778.0	10.0	742.5	-	1029
Vapor boilup (kmol/h)	115.8	1023	-	738.6	1343
Cooling duty (GJ/h)	34.8	7.22	33.4	-	50.8
Reboiler duty (GJ/h)	4.42	37.7	-	26.3	51.0
Preheater (GJ/h)		0.8		15.5	
Composition (mol frac.)					
Feed	0.5/0.5	0.5/0.5	0.5/0.5	0.5/0.5	0.5/0.5
Product-methanol	0.950	0.950	0.972	0.928	0.950
-ethanol	0.947	0.951	0.970	0.934	0.950

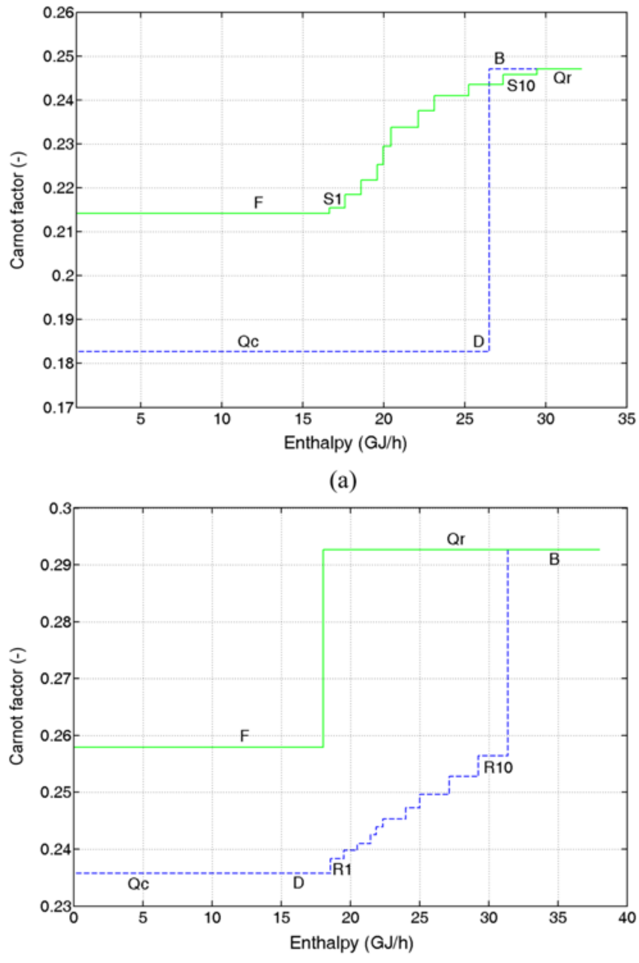
umn cannot be recycled to heat its own stripping section due to the negative temperature difference. When two distillation columns process the feed of a single binary column in a half each, the necessary temperature difference can be generated by applying different pressures to the columns. The performance of energy saving and the thermal efficiency is examined and compared with that of the conventional distillation system. Table 1 shows the results of structural design and operating condition computation for the example process of the benzene-toluene mixture. The total number of trays of the conventional system is taken as the sum of tray numbers of the two columns in the proposed system. The lower ten trays of the first column in the proposed system and the upper ten trays are internally heat integrated. The temperatures of the lower ten trays of the first column are between 94.8 °C and 109.6 °C, while the upper ten trays of the second are between 105.9 °C and 115.1 °C. Douglas [31] suggested a minimum approach temperature of 10 °F in the heat exchanger design, and a smaller temperature difference means a lower operating pressure at the rectifying section [32].

The comparison of energy demand indicates that the proposed system requires 17.4% less of heating duty and 22.5% less of cooling duty. As expected, a significant reduction of heat duty is observed at the reboiler of the first column and the condenser of the second, where internal heat integration is applied. The internal heat integration in the ternary separation system of BTX mixture [8] reduces the reboiler duty by 17.1% and condenser duty by 25.9%, which is comparable to the present study. These reductions also save the investment cost of the heat exchanger fabrication generally much higher than the construction cost of column section [30,33]. The reduction of heating duty is less than one-third of the predicted reduction of 60% in the ideal HIDiC [24], and it is because heat integration is applied to half of the two columns. Note that the ideal HIDiC consumes electric energy. The heating duty of this study includes the preheater demand for the feed temperature increase in the second column. The temperature of the second column in the proposed system is higher than the feed temperature, and so the preheater is installed for the temperature adjustment.

The comparison results of exergy loss and thermal efficiency between the proposed and conventional systems in the benzene-toluene process are listed in Table 3. The exergy loss was calculated from the enthalpy-Carnot factor diagram shown in Fig. 5. The top demonstrates for the first column of the heat-integrated distillation

Table 3. Exergy losses in the proposed, existing double effect and conventional distillation systems for the benzene-toluene process. Units are in GJ h^{-1}

Process	Proposed		Double effect		Conventional
	1st	2nd	1st	2nd	
Trays	0.518	0.251	0.687	1.032	0.149
Condenser	0.485	0.150	0.503	–0.212	0.897
Reboiler	0	0.599	0.023	0	1.228
Preheater		0.296		4.502	
Total		2.299		6.534	2.274
Feed	0.828		0.828		0.828
Overhead	0.121	0.253	0.118	1.793	0.241
Bottom	0.339	0.562	0.322	0.716	0.728
Min. work		0.447		2.120	0.141
Thermal Efficiency (%)		16.3		24.5	5.8



(a)
(b)

Fig. 5. Enthalpy-Carnot factor diagram of the heat-integrated distillation in the benzene-toluene process: (a) 1st column (b) 2nd column.

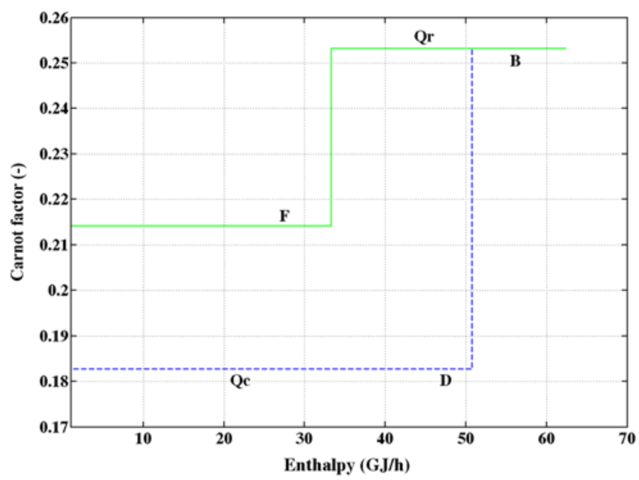


Fig. 6. Enthalpy-Carnot factor diagram of the conventional distillation in the benzene-toluene process.

system, and the bottom does for the second. The diagram for the conventional distillation column is given in Fig. 6. The exergy loss is calculated as the area between the supplied exergy lines on top lines and the recovered exergy lines of bottom. Though the total enthalpy difference in the conventional system is much larger than that in the proposed, the exergy losses of two systems are nearly identical. It is because the difference of the Carnot factor for the proposed system is much larger than that of the conventional. For heat integration the pressure of the second column was raised to give enough temperature difference between the second and first columns. The increased column temperature gives the high Carnot factor. The temperature elevation also increases the minimum work calculated from Eq. (6) of the proposed system. The exergy values of feed and products are obtained from the HYSYS simulation. The thermal efficiency computed from Eq. (5) of the proposed system is almost three-times that of the conventional system. The tray-by-tray heat integration is responsible for the increase in efficiency. Because the reboiler temperature found from the HYSYS simulation is close to the temperature of bottom product and the enthalpy of the product is much larger than the reboiler duty of the first column as shown in the top plot of Fig. 5, the exergy loss in the reboiler becomes zero.

Table 2 shows the results of structural design and operating condition computation for the methanol-ethanol process. The proposed system utilizes two 24-tray columns instead of the 20-tray columns used in the benzene-toluene process. The initial computation indicated that the total number of trays of 20 is close to the minimum tray number requiring a large reflux flow. Therefore, the tray number was raised to 24. The lower 12 tray of the first column in the proposed system and the upper 12 trays are internally heat integrated. The temperatures of the lower 12 trays of the first column range over 73.4 °C and 78.8 °C, while the upper 12 trays of the second do over 80.3 °C and 85.9 °C. In the energy duty comparison, the proposed system requires 15.8% less of heating duty and 17.3% less of cooling duty. These savings are little less than the benzene-toluene process. While the exergy losses are nearly the same in both systems, the thermal efficiency of the proposed system is much higher than the conventional system as listed in Table 4.

Table 4. Exergy losses in the proposed, existing double effect and conventional distillation systems for the methanol-ethanol process. Units are in GJ h^{-1}

Process	Proposed		Double effect		Conventional
	1st	2nd	1st	2nd	
Trays	−0.494	−0.927	−0647	−0.741	−0.746
Condenser	−0.703	−0.101	0.694	−0.261	−1.027
Reboiler	0	0	0.120	0	0
Preheater		−0.354		3.548	
Total		−2.579		2.714	−2.519
Feed	−4.009		−4.009		−4.009
Overhead	−0.788	−1.300	−0.782	0.600	−1.575
Bottom	−1.140	−1.585	−1.068	−1.926	−2.428
Min. work		−0.804		0.833	−0.006
Thermal Efficiency (%)		23.8		23.5	0.2

The performance comparison of the proposed system with the double-effect distillation column shown in Fig. 2 indicates that 3.5% of the heating duty is reduced in the benzene-toluene process as listed in Table 1. In the methanol-ethanol process 2.7% more heating duty is required in the proposed system as found in Table 2. However, the operating pressure of the second column in the double-effect system is 50% to 67% higher than that of the proposed system, and therefore utility cost is more with the double-effect distillation system due to the high pressure steam used in the high pressure column. For thermal efficiency, the double-effect distillation system gives higher efficiency in the benzene-toluene process, but a little lower efficiency is yielded in the methanol-ethanol process.

For the economic evaluation of the proposed distillation column, the investment and operating costs were calculated using the following equations. The evaluation of column cost is given as

$$C_{col}=\left(\frac{M\&S}{280}\right)C_f D_C^{1.066} H_C^{0.802} C_p \tag{7}$$

where the M&S is the Marshall and Swift index and the value of the third quarter of 2,010 of 1,473.3 is used here. The pressure related cost coefficient C_f is given in Olujic et al. [34]. The column diameter is found from

$$D_c=0.08318\sqrt{V} \tag{8}$$

where V is the rate of vapor in kg-mol/h. The height of column is calculated from two foot spacing and the total number of trays.

The cost of trays is

$$C_{tray}=\left(\frac{M\&S}{280}\right)97.243 D_C^{1.55} H_C F_c C_p \tag{9}$$

where the fabrication coefficient F_c is given in Olujic et al. [34]. The penalty coefficient C_p of 1.2 was used for the HIDiC. The cost of the condenser is

$$C_{cond}=\left(\frac{M\&S}{280}\right)1609.13 A_C^{0.65} \tag{10}$$

where A_C is the heat transfer area of the condenser. Similarly, the cost of the reboiler is

$$C_{reb}=\left(\frac{M\&S}{280}\right)1775.26 A_R^{0.65} \tag{11}$$

The steam and cooling costs are \$13 per ton and \$0.03 per ton, respectively. The total annual cost (TAC) was calculated from the investment cost of column, trays and heat exchangers, and the operating cost includes steam and cooling water costs. The investment cost is counted with the 10 year payback. Table 5 shows the cost comparison between the proposed and conventional systems. The investment cost of column and heat exchanger construction of complete fabrication and utility cost was examined. The proposed system requires 16.8% less total annual cost than the conventional distillation column in the benzene-toluene process and 12.3% less in the methanol-ethanol process.

A key advantage of the proposed distillation system over the HIDiC is that it does not employ a compressor that is difficult to operate and to maintain. By applying a heat-integrated column between two conventional columns, more than 15% energy saving is yielded as shown in the two example processes. The role of the heat-inte-

Table 5. Economic evaluation of the proposed heat-integrated and conventional binary distillation systems for the benzene-toluene and methanol-ethanol processes. Units are in million U.S. dollars

Process	Benzene-toluene			Methanol-ethanol		
	Proposed		Conventional	Proposed		Conventional
	1st	2nd		1st	2nd	
Investment						
Column	0.356	0.402	0.700	0.590	0.604	0.906
Tray	0.027	0.033	0.071	0.055	0.057	0.103
Heat exchanger	0.353	0.520	0.838	0.517	0.650	1.107
Utilities						
Steam	0.122	0.931	1.307	0.172	1.519	1.994
Cooling water	0.023	0.011	0.043	0.045	0.010	0.066
Total annual cost	0.219	1.038	1.511	0.333	1.660	2.272

grated column is like an economizer of a distillation column. The use of the compressor in the HIDiC hampers its use in chemical processes [3]. Though the proposed system does not use a compressor, the operating pressure of the second column is higher than that of the first to have enough temperature difference required for internal heat integration. The pressure elevation in the binary separation of this study is higher than that in the ternary BTX separation [8]. In the case of ternary separation the processing material of the second column contains heavy components, but the materials of both columns in the binary separation of this study are the same. Therefore, a larger pressure difference is necessary to give the temperature difference for the heat integration between the two columns.

The proposed distillation system in this study utilizes separate condensers and reboilers in the columns, which helps the column operation. Though the two columns are thermally connected, the separate manipulation of the four heat exchangers gives easy control of the pressure and temperature of the columns. The column pressure can be adjusted with the condenser duty, and the individual column is operated with a separate condenser. Whereas, the temperature at the top tray is controlled with the reflux flow rate, and that of the bottom is done with the reboiler duty in the same manner as in the control of a conventional distillation column. Therefore, the operation of the proposed system is as simple as that of a conventional column.

CONCLUSIONS

An internally heat-integrated distillation column is proposed for double-effect distillation, and its energy savings and thermal efficiency are compared with those of the conventional distillation column. Using two identical binary columns, the heat integrated system for a ternary separation developed previously has been applied to the binary separation. Two examples of the benzene-toluene and methanol-ethanol processes are used for the performance evaluation. The performance comparison indicates that the proposed system requires 17.4% less reboiler duty and 22.5% less condenser duty for the benzene-toluene process and 15.8% less heating duty and 17.3% less cooling duty for the methanol-ethanol process. The thermal efficiencies are 16.3% and 23.8% for the benzene-toluene and methanol-ethanol processes, respectively, which are much higher than those of the conventional distillation system. In the proposed system no compressor is utilized to eliminate the problem caused by the compressor, and the separate reboilers and condensers for each column give easier operation than the original internally heat-integrated distillation column.

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NOMENCLATURE

B : bottom product flow rate [kmol/h]
D : overhead product flow rate [mol/h]
E : exergy [kJ/kmol]
F : feed flow rate [kmol/h]
f : carnot factor [-]
H : enthalpy [kJ/kmol]
L : liquid flow rate [kmol/h]
Q : heat transfer rate [kJ/h]
R : tray heat removal rate [kJ/h]
S : entropy [kJ/kmol K] or tray heat supply rate [kJ/h]
T : absolute temperature [K]
V : vapor flow rate [kmol/h]
W : work [kJ]

Superscripts
B : bottom
D : overhead
F : feed
L : liquid
V : vapor

Subscripts
c : condenser
j : tray number
m : minimum
o : standard
Q : energy
r : reboiler

Greek Letters

η : thermal efficiency

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