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DESIGN OF HEAT INTEGRATED DISTILLATION SYSTEMS FOR A LIGHT ENDS SEPARATION

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ABSTRACT

This paper presents an industrial case-study: the synthesis of heat-integrated distillation systems applied to the light ends separation section of olefin and gas process plants. The distillation systems presented in this work employ heat-integration principles to significantly reduce the heat requirements compared to the traditional simple column train. The work started from the simulation of the existing plant, by which the parameters of the system were identified. Then the possible sequences of simple columns with sharp splits were identified for the considered application, and all columns of the configurations were designed and the heat-integrated configuration was also considered. In order to verify the examined distillation systems, all simple and complex configurations were simulated by rigorous numerical models. On the basis of the numerical simulations, the energy requirements for each configuration were evaluated. A rating of different plants was then performed, based on the total annual cost, allowing identifying the best plant configuration. A heat-integrated configuration showed the best performances for the considered separation

INTRODUCTION

Distillation is the most widely used separation technique in the petrochemical and chemical process industries for the separation of fluid mixtures despite its high energy requirement. Significant energy saving can be reached by the use of complex distillation arrangements such as the side-stripper, the side-rectifier, the thermal (internal) column coupling (also known as petlyuk system), the (external) energy integration (also known as energy integrated distillation system) and the heat pumping techniques [1-3]. The energy saving in distillation is an essential field of the chemical engineering research since the early 1970; appropriate integration of the distillation columns with the overall process often results in a significant reduction of the energy

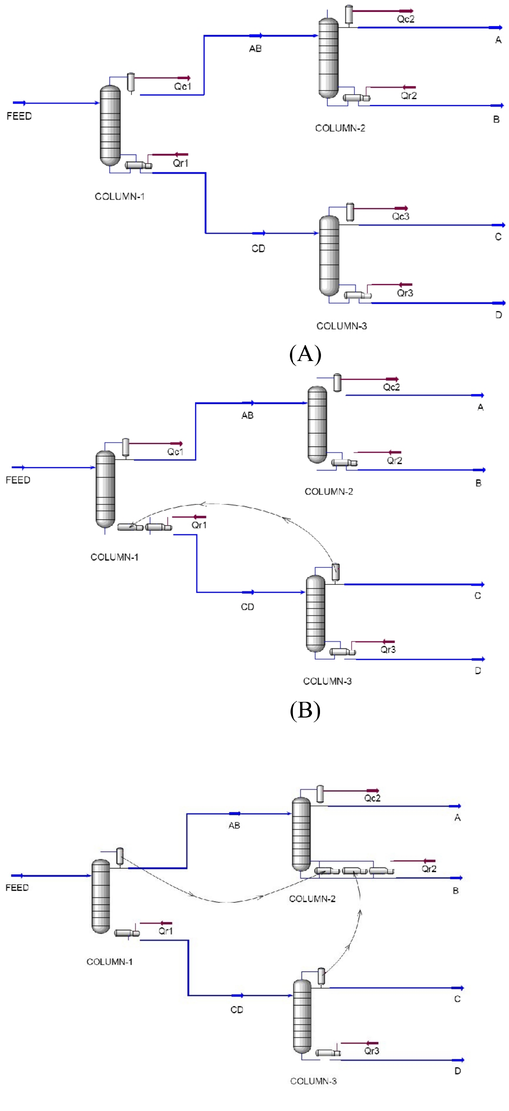
NOMENCLATURE

S_n	[-]	number of all simple configurations
n	[-]	number of components
$M\,\&\,S$	[-]	Marshal and swift index
D	[m]	Diameter of column
A	[m^2]	Area of exchanger
H	[m]	Height of column

Consumption in many cases. The classical design of a Nomenclature distillation system for a multicomponent separation uses only the simple columns. Each of the simple columns in a multicomponent distillation configuration receives a feed and performs a sharp split between two adjacent components of the feed mixture. Meanwhile, each of the simple columns produces a top product with a condenser and a bottom product with a reboiler. Thus, in any case of the simple column configuration for an n-component mixture, it needs n - 1 simple columns with n - 1 condensers and n - 1 reboilers [4, 5]. The simple column configurations for multicomponent distillation are simple and easy to design and operate. Because the number of separation sequences increases dramatically when the number of components in the feed mixture increases, a considerable number of works have been conducted on optimal synthesis of simple column configurations for an n-component distillation [4-7]. Thompson and King reported the following formula to calculate the number of all simple configurations [8]:

$$S_n = \frac{[2(n-1)]!}{n!(n-1)!} \qquad (1)$$

The task for the design of a heat integrated simple column configuration consists of searching for the possible heat matches among all of the condensers and reboilers of the simple columns. Large numbers of works have been done for the synthesis of the optimal heat integrated simple column configurations since the work by Rathore et al [9-16]. This paper considers the heat integrated distillation systems for a chemical process in order to minimize the total annual costs. The plant configuration considered in our case shown in (figure 1-A). This configuration can be classified as a distributed sequence and can be represented by the following separation task: AB/CD; A/B; C/D. Other four simple column configurations are possible for this separation, the direct (A/BCD; B/CD; C/D) sequences (Figure 2-A) and the direct-indirect (A/BCD; BC/D; B/C) sequences (Figure 3-A), the indirect (ABC/D; AB/C; A/B) sequences (Figure 4-A) and the indirect-direct (ABC/D; A/BC; B/C) sequence (Figure 5).

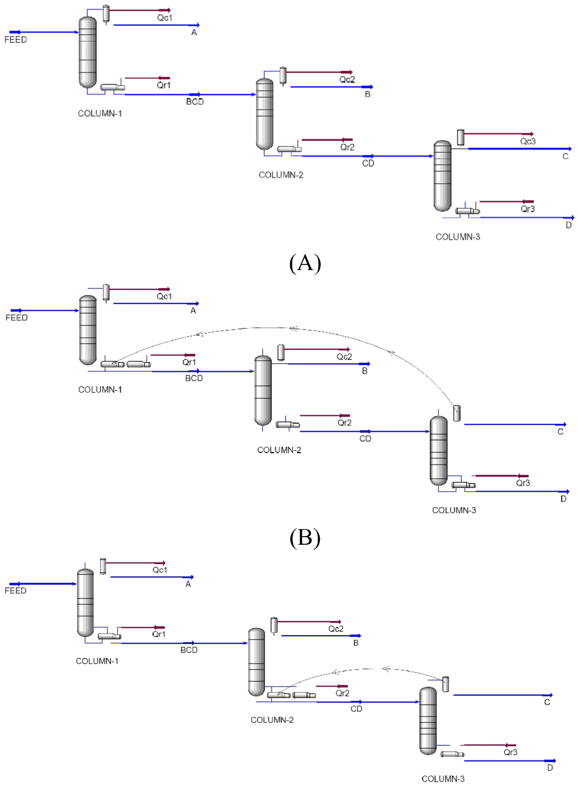


(A)

(B)

(C)

Fig. 1. Distributed sequences

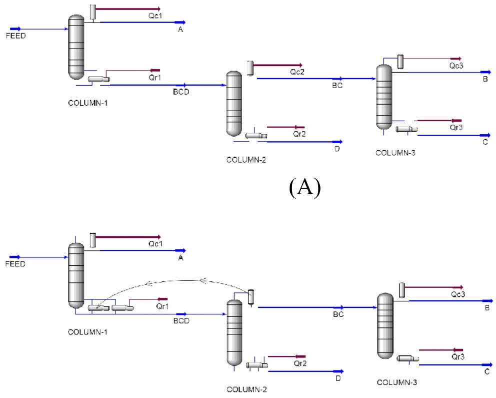


(A)

(B)

(C)

Fig. 2. Direct sequences



(A)

(B)

Fig. 3. Direct-Indirect sequences

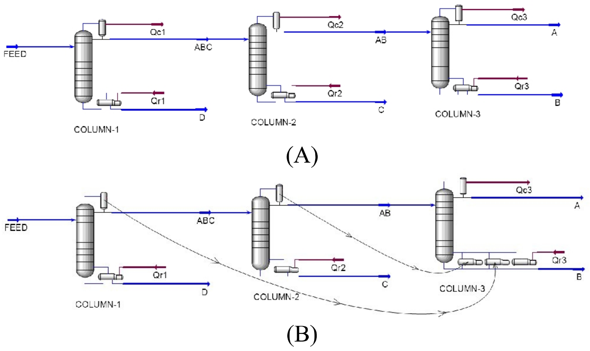


Fig. 4. Indirect sequences

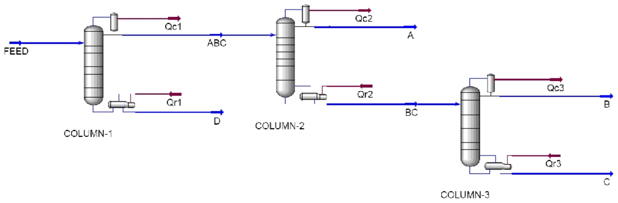


Fig. 5. Indirect-Direct sequence

CASE STUDY

The process examined in this paper, is the separation of a multi-component flow to four categories of products with a purity of 0.99. The intake flow has a temperature of 59.3 ^{o}C, a pressure of 1785 kPa and a flow rate equal to 1355 kmole/hr. Compositions of process intake flow and products are shown in table 1. The initial process is illustrated in figure 1-A and it is regarded as a reference case for other states. Simulations are carried out using the software HISYS and two models, Shortcut and Rigorous, are employed for simulating the towers. The governing equations for simulating the columns are NRTL equations.

Product groups	Feed Components	Feed Xi (Mol Fraction)
A	C_2H_4	0.0001
	C_3H_4	0.0047
	C_3H_6	0.4538
B	C_3H_8	0.2912
	C_4H_4	0.0003
C	C_4H_6	0.0011
	C_4H_8	0.1541
	C_4H_{10}	0.0359
D	C_5H_{12}	0.0419
	C_6H_{14}	0.0137
	C_7H_{16}	0.0031

Table 1- Feed and product specifications

The TAC function is optimized using the optimizer of HYSYS software. The method SQP is used to minimize TAC function. Different arrangements have been obtained for the process with heat integration satisfying the constraints of the problem. These arrangements are shown in figures 1-11. In this paper, first all possible states for arrangement of columns in separation of the process intake flow are simulated and then attempts are made for optimization and heat integration of columns. Utility costs are shown in table 2[16]. Equations related to installation and start up costs of columns is provided in appendix A [17]. Results of design and optimization of arrangements are shown in table 3.

Utility	Temperature Level (° C)	Values
LP-steam ($/ton)	158	13
MP-steam ($/ton)	200	16
HP-steam ($/ton)	250	20
Cooling water ($/ton)	35-45	0.082

Table. 2. Utility costs

Description	Primary Distributed			Distributed (C3-R1)		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3
Pressure (kpa)	1785	1940	735	1901	2217	2748
Diameter (m)	3.35	4.88	1.67	3.35	4.89	1.74
Reflux ratio	0.71	8.45	0.64	0.72	8.53	0.79
Actual plates	54	163	43	54	166	42
Total actual plates	260			262		
Heating rate (kJ/h)	111423972			106272677		
Cooling rate (kJ/h)	117477072			107204581		
HX duty (kJ/h)	-			7314222.4		
Steam cost ($/yr)	4535992.7			4557609.8		
C.W cost ($/yr)	1843659.3			1682445.1		
Operat. cost ($/yr)	6379652			6240054.9		
Capital cost ($/yr)	758906.59			555824.96		
TAC ($/yr)	7138558.6			6795879.9		
Op. cost saving (%)	-			2.19		
Capital saving (%)	-			26.76		
TAC saving (%)	-			4.8		

Table. 3. A- Optimal schemes for Primary Distributed and Distributed (C3-R1)

This table shows that the arrangement Distributed (C1&C3-R2) (figure 1-C) with a TAC optimization of 27.86% represents a better result in comparison to other states. Arrangements Indirect(C1&C2-R3), with a TAC optimization of 17.15%, Direct(C3-R1) with an optimization of 13.14%, Direct(C3-R2) with an optimization of 11.1%, Direct with an optimization of 9.85%, Direct-Indirect(C2-R1) with an optimization of 6.48%, Distributed(C3-R1) with an optimization of 4.8%, Direct-Indirect with an optimization of 1.2% respectively, represented better result compared to the initial arrangement. The arrangements Indirect and Indirect-Direct increased the TAC 17.31% and 8.11% respectively.

Description	Distributed (C1&C3-R2)			Direct		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3
Pressure (kpa)	2989	2220	1042	2365	2113	651
Diameter (m)	3.41	4.92	1.72	6.29	2.39	1.68
Reflux ratio	0.77	8.62	0.74	8.95	1.21	0.65
Actual plates	43	167	44	176	42	41
Total actual plates	254			259		
Heating rate (kJ/h)	78498124			100110119		
Cooling rate (kJ/h)	82738403			105533811		
HX duty (kJ/h)	32886838			-		
Steam cost ($/yr)	3326005			4224193.8		
C.W cost ($/yr)	1298478.3			1656224.4		
Operat. cost ($/yr)	4624483.2			5880418.2		
Capital cost ($/yr)	525335.21			554833.01		
TAC ($/yr)	5149818.4			6435251.2		
Op. cost saving (%)	27.51			7.83		
Capital saving (%)	30.78			26.89		
TAC saving (%)	27.86			9.85		

Table 3. B- Optimal schemes for Distributed (C1&C3-R2) and Direct

Description	Direct (C3-R1)			Direct (C3-R2)		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3
Pressure (kpa)	2414	1969	1698	2313	1839	2667
Diameter (m)	6.29	2.38	1.71	6.28	2.37	1.74
Reflux ratio	8.97	1.2	0.73	8.93	1.18	0.78
Actual plates	174	40	44	176	40	49
Total actual plates	258			265		
Heating rate (kJ/h)	94211676			96031322		
Cooling rate (kJ/h)	96995376			97473954		
HX duty (kJ/h)	7832094			7279755.3		
Steam cost ($/yr)	4127335.1			4238147.6		
C.W cost ($/yr)	1522224.1			1529734.8		
Operat. cost ($/yr)	5649559.2			5767882.4		
Capital cost ($/yr)	550778.59			578258.09		
TAC ($/yr)	6200337.8			6346140.5		
Op. cost saving (%)	11.44			9.59		
Capital saving (%)	27.42			23.8		
TAC saving (%)	13.14			11.1		

Table 3 C- Optimal schemes for Direct (C3-R1) and Direct (C3-R2)

Description	Direct-Indirect			Direct-Indirect (C2-R1)		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3
Pressure (kpa)	2080	1307	1854	2024	2740	1717
Diameter (m)	6.27	2.37	1.83	6.27	2.42	1.9
Reflux ratio	8.9	0.31	1	8.89	0.37	1.13
Actual plates	174	44	42	174	49	46
Total actual plates	260			269		
Heating rate (kJ/h)	109826812			96202845		
Cooling rate (kJ/h)	113397355			99389758		
HX duty (kJ/h)	-			14648275		
Steam cost ($/yr)	4634066			4414348.8		
C.W cost ($/yr)	1779633.1			1559801		
Operat. cost ($/yr)	6413699.2			5974149.8		
Capital cost ($/yr)	639173.17			701488.26		
TAC ($/yr)	7052872.3			6675638		
Op. cost saving (%)	-0.53			6.36		
Capital saving (%)	15.78			7.57		
TAC saving (%)	1.2			6.48		

Table 3. D- Optimal schemes for Direct-Indirect and Direct-Indirect (C2-R1)

Description	Indirect (C1&C2-R3)			Indirect-Direct		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3
Pressure (kpa)	2585	2986	2169	1844	2296	1912
Diameter (m)	3.2	2.71	4.92	3.18	5.01	1.89
Reflux ratio	0.25	0.72	8.61	0.22	8.69	1.12
Actual plates	51	49	168	48	167	44
Total actual plates	268			259		
Heating rate (kJ/h)	82072805			117517206		
Cooling rate (kJ/h)	83029960			120136244		
HX duty (kJ/h)	48112352			-		
Steam cost ($/yr)	4003977.4			5307057.3		
C.W cost ($/yr)	1303053.9			1885391.8		
Operat. cost ($/yr)	5307031.3			7192449		
Capital cost ($/yr)	607282.93			525144		
TAC ($/yr)	5914314.3			7717593		
Op. cost saving (%)	16.81			-12.74		
Capital saving (%)	19.98			30.8		
TAC saving (%)	17.15			-8.11		

Table 3 E- Optimal schemes for Indirect (C1&C2-R3) and Indirect-Direct

The Distributed arrangement of the initial process is considered as the reference of comparison for other arrangements (Figure 1-A)

In the arrangement Distributed (C1&C3-R2), as can be seen in table 3 and figure 2, by increasing pressures of columns 1 and 3 and creating the proper temperature difference, the heat output of these towers is transferred to the reboiler of column 2 and thus cooling costs of condensers 1 and 3 are eliminated and the heating of the reboiler of column 2 is considerably reduced. This has resulted in reduction of operational costs to an amount of 27.51% compared to the reference case. Comparing diagrams 12 and 13, illustrates the degree of heat integration in 62.86°C and temperature change of the columns relative to the reference state. In this state, the reboilers of all towers are fed with LP steam which lowers the cost of consumption of steam in reboilers. Moreover, new designs of columns and heat exchangers have decreased initial costs to an amount of 30.78%. In the end, TAC has decreased by 27.86%. Among the reasons involving in high efficiency of this arrangement is separation of products of groups A and B in tower number 2, and away from the products of groups C and D. Products A and B are primarily propane and propylene. Because their relative volatilities are close to each other, their separation is difficult. Their molar percentages while entering the process is high compared to other compounds which have caused the heat consumption, height and diameter of column 2 to be higher than other columns in the process. One can say that in this arrangement, as a result of separation of products A and B in one column and also in the intermediate columns of the process, better conditions have been provided for the possibility of better heat integration and lower annual costs.

CONCLUSIONS

The obtained results show that the best arrangements were the ones in which two products A and B with high and close –to-each other values of relative volatility were separated in one column. On the other hand, because their molar percentages are high compared to other products, this column is preferred to be one of the primary or intermediate columns. For example in the state Indirect(C1&C2-R3), since products A and B are separated in column 3 and are present with high molar percentages in columns 1 and 2, the heat consumption of these columns have become more than that of columns 1 and 3 in the case Distributed(C1&C3-R2) and consequently the operational costs have increased. Because products A and B are separated in one column, only in cases Distributed(C1&C3-R2) and Indirect(C1&C2-R3) one can bring about two different heat integrations between condensers and reboilers of different columns. Low temperature difference inside the columns increases the freedom in changing columns' temperatures and heat integration of the column. The temperature difference in one column depends on the value of volatility of the components inside that column and molar percentages of these components in the intake flow of the column. Since molar percentages of components A and B are high, being respectively 45% and 29% of the process intake flow and also because of the proximity of their relative volatilities, the temperature difference of the columns which separate A from BCD and the columns which separate A from B are low. It can be deduced that in order to select arrangements with low temperature difference in columns, relative volatilities and molar percentages of the components should be investigated from the very beginning.

From table 3 it can be seen that the sum of costs related to consumption of steam and cold water in TAC has an order of 10^6 while the sum of initial costs of columns and heat exchangers is of the order 10^5. Consequently, it can be said that in designing the arrangements, one needs to increase the number of heat integrations and the amount of transferred heat since the amount of decrease of operational costs will be considerably higher than the increase of initial costs. Therefore, the chosen arrangements should be the ones which provide the possibility of more heat integration in the process.

In table 3 it can be seen that the sum of costs of the cold water is a number of the order 10^5 but the sum of the costs of the steam is a number of the order 10^6. This is to say that more attention should be paid to temperature reduction and heat consumption of reboilers in the optimization procedure of columns.

By examining the initial costs we can come up with this conclusion that it is better to keep a higher value for the minimum temperature difference in the condensers of the columns which exchange a large amount of heat with cold water, so that their finished costs are lowered. The obtained results can be summarized as follows:

1- Choosing arrangements which provide the possibility for largest amounts of integration with high heat transfer potential
2- Separation of components with close volatilities in one column and away from the other components
3- Separation of components with high molar percentages in primary or intermediate columns
4- Keeping the heat of reboilers lower, and lowering consumption of expensive MP and HP steams
5- Investigating the values of volatilities and molar percentages of the components of the intake flow in order to choose arrangements with lower temperature differences in columns
6- Separation of ultra light and ultra heavy components in separate columns in order to lower the temperature difference between the condenser and the reboiler of the column.
7- Keeping higher the minimum temperature difference in condensers which have high heat loads.

The optimum state can be easily found considering the above conditions and that way, many of the arrangements will be eliminated in the primary stages of optimization of columns.

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APPENDIX A. COST ESTIMATION

A.1. Distillation columns

The total cost of a distillation column can be considered as a sum of the costs of column shell and trays. The number of stages and the diameter values obtained from the Aspen simulations can be utilized in the following correlations. All the correlations are valid for carbon steel construction and are updated from mid-1968 to 2009 utilizing the Marshall and Swift cost index [18].

• Column shell

$$\$ = \left(\frac{M \& S}{280}\right)(937.61)D^{1.066}H^{0.802}$$

(A.1)

$$H = (N-1)\times 0.6 + 6.0$$

(A.2)

The correlation (A.1) is valid for a pressure less than 345 kPa; otherwise a correction factor must be applied. The column height is evaluated considering 0.6m as tray spacing and 6.0m as disengagement.

• Column trays

Considering sieve trays, the following correlation has been utilized.

$$\$ = \left(\frac{M \& S}{280}\right)(97.24)D\ H^{0.802}$$

(A.3)

A.2. Heat exchangers

The heat exchanger cost evaluation is based on the heat exchanger area evaluated utilizing the usual design formula:

$$A = \frac{Q}{U\ \Delta T_M}$$

(A.4)

Mean values of 1800 kJ/ (m2 h ◦C) and 2100 kJ/(m2 h ◦C) for the overall heat transfer coefficient for condensers and reboilers are assumed respectively. Assuming shell and tube, floating head and carbon steel construction the cost correlation is as follows:

$$\$ = \left(\frac{M \& S}{280}\right)(474.67)A^{0.65}$$

(A.5)

The reported correlation is valid for design pressure less than 1034.2 kPa and an exchange area range between $18.6 < A < 464.5$m2.

A.3. Annual capital cost

The capital cost (purchase plus installation cost) is annualized over a period which is often referred to as plant life time.

Annual capital cost = Capital cost / Plant life time

TAC = Annual operating cost + Annual capital cost

Operating cost was assumed just utility cost (steam and cooling water).

- Plant lifetime = 10 years
- Operating hours = 8000 h/year