Gas Processing Journal

Vol. 1 No. 1, January 2013
ISSN (Print): 2322-3251
PP. 51-68

The Design and Optimization of Distillation Column with Heat and Power Integrated Systems

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(Received: 1September 2012, Accepted: 4 December 2012)

Abstract: *Based on two integration steps, an optimization framework is proposed in this work for the synthesis and design of complex distillation sequence. The first step is to employ heat integration in sequence and reduce the heat consumption and total annual cost of the process. The second one is to increase the exergetic efficiency of sequence by generating power in implemented expanders in sequence. The profit of power generation directly affects the operating cost of the process and decreases the total annual cost. In each step, the target is to minimize the objective function of total annual cost. A simulator is used to simulate the equipment's specification and formulate the objective function of cost. Results from employing these two integration steps for the considered case study show the advantages of such a complex distillation sequence with heat integration and power generation. The results represent a very high improvement for the sequence Indirect since the properties of the intake flow to the process are in a way that in this sequence not only do we have a high freedom for carrying out heat integration, but a large amount of power is also produced between the columns due to having high flow rate flows between the columns.*
Keywords: *Distillation, Sequence, Modeling, Integration, Optimization, Expander*

1. Introduction

Using simple distillation columns, the separation of the components of multi-component flows can be carried out by using different sequences. Since distillation columns are very energy consumptive, and different sequences have different values of energy requirement, choosing the best sequence is economically important. On the other hand heat integration

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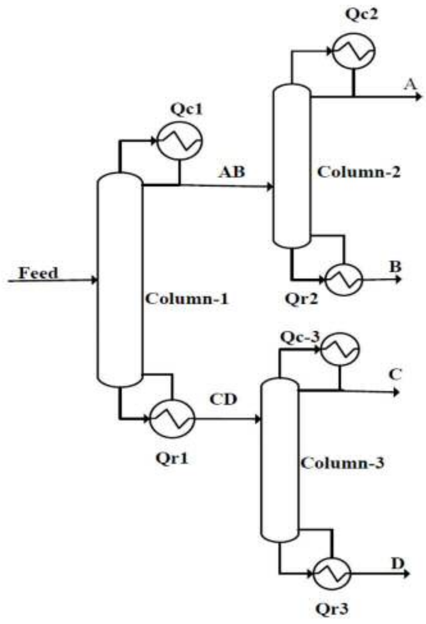
causes a considerable reduction in heat consumption of these column sequences (Proios et al. 2005). There are a number of different methods for heat integration of columns, and numerous works have been carried out about this subject so far (Andrecovich& Westerberg 1985; Kattan & Douglas 1986; Smith & Linnhoff 1988; Isla & Cerda 1988; Trigueros & Velasco 1989; Agrawal 1996; Yeomans & Grossman 1999; Sobocan & Glavic 2000; Proios et al. 2005).

Another possible way for optimizing the columns is to increase the possibility of power production between columns. In processes such as separation systems of olefin units in which some of columns operate in temperatures lower than that of the environment and exchange heat using cryogenic systems, the power produced between high temperature columns can be directly consumed by cryogenic systems. Power generation in column sequences can highly decrease the heat and power consumptions of the process, and increase the thermodynamic efficiency of separation systems. The cost analysis of the separation system shows the effect of power generation on the cost function of the process. Exergy analysis can also display the effects of power generation on the performance of separation systems. Numerous works of research have been done so far related to the exergetic examination of columns and sequences (Douani et al. 2007; Araújo et al. 2007; Kistar, 1992; King, 1980; Errico et al. 2009).

Most of these analyses allow the evaluation of the exergy losses and exergetical efficiencies to identify potential locations for process improvements. From the thermodynamic point of view, simple sequences of columns have lower exergetic efficiencies, for instance the efficiency of the direct sequence is usually 10% (Chemical Engineering Magazine, 2009). Using the capability of power production between the columns greatly increases the exergetic efficiency of the process.

In this paper, separation of a multicomponent stream into four groups of products is examined. In order to separate multicomponent flows to n product groups, the number of column sequences (S_n) is calculated using King's equation (Mascia et al. 2007):

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



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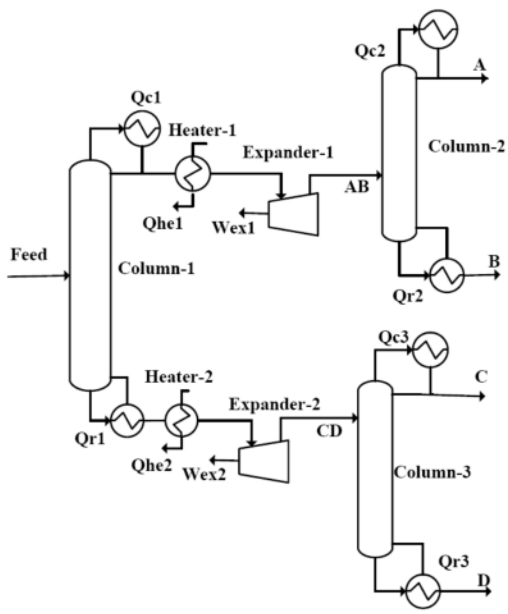

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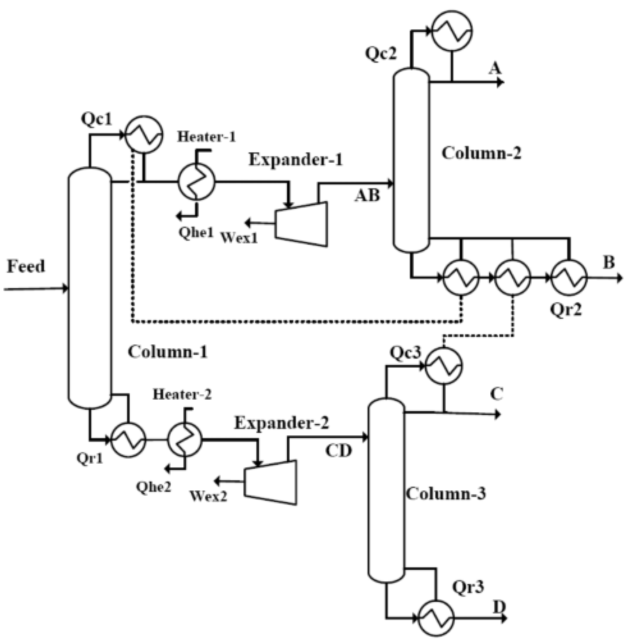

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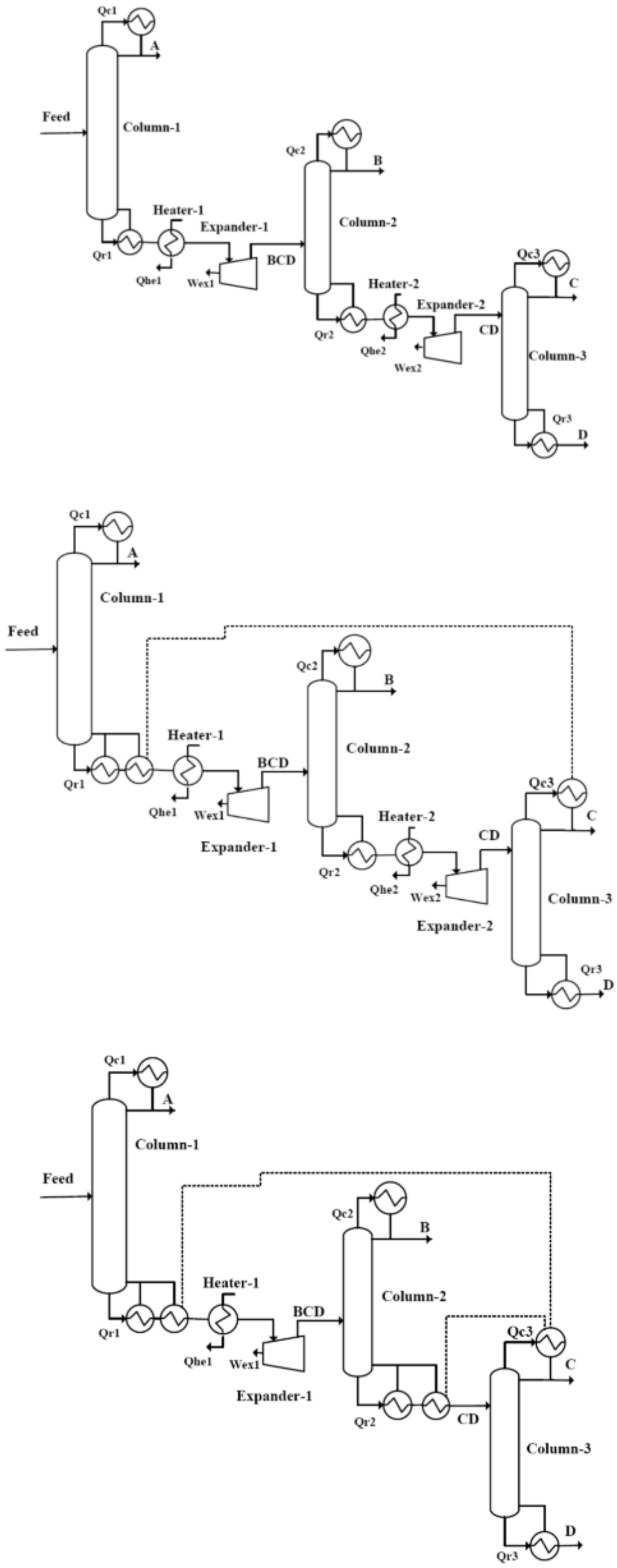
Fig. 1. Distributed sequences: A is Primary Distributed, B is Distributed(1X2&1X3), and C is Distributed(C1&C3-R2)(1X2&1X3

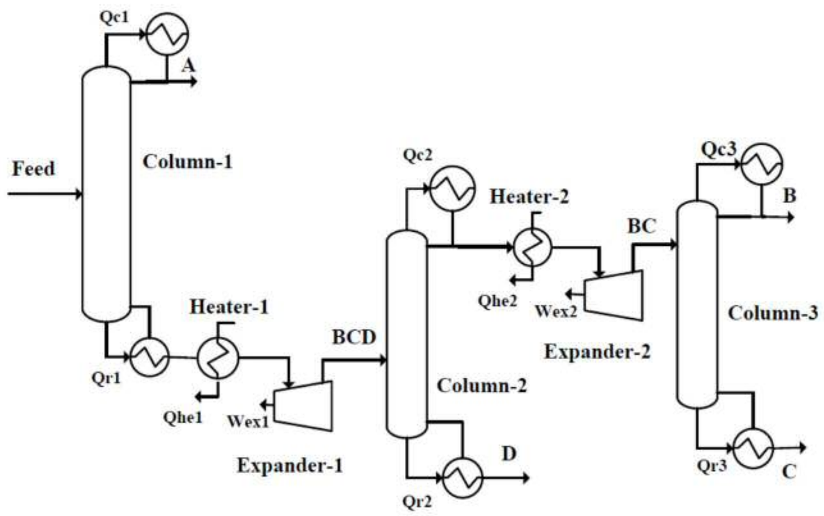


(A)

(B)

(C)

Fig. 2. Direct sequences: A is Indirect(1X2&2X3), B is Direct(C3-R1)(1X2&2X3), and C is Direct(C3-R2&R1)(1X2)



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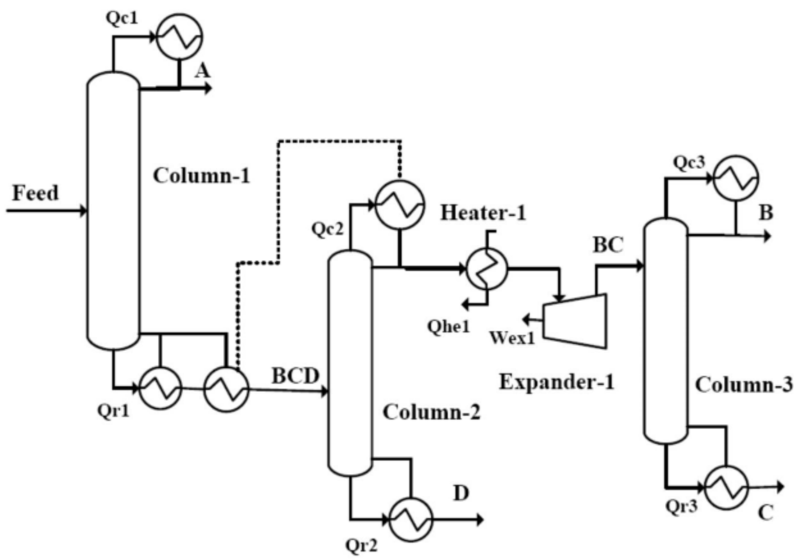

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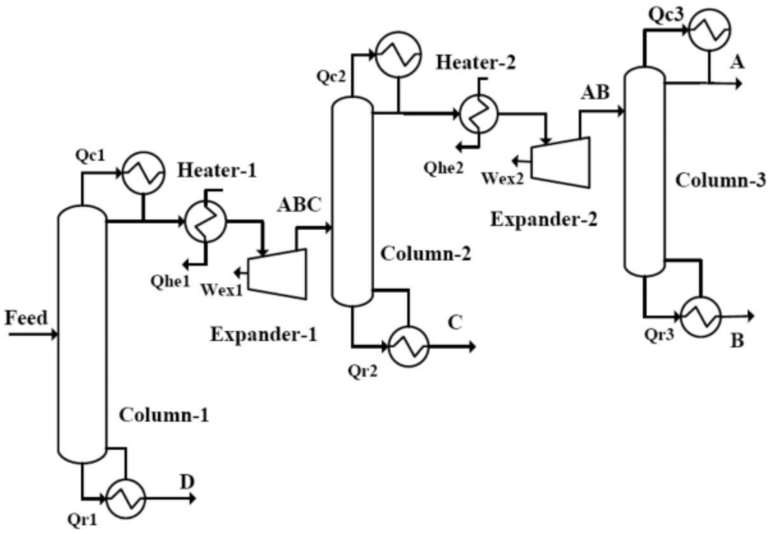
Fig. 3. Direct-Indirect sequences: A is Direct-Indirect(1X2&2X3),B is Direct-Indirect(C2-R1)(2X3)



(A)

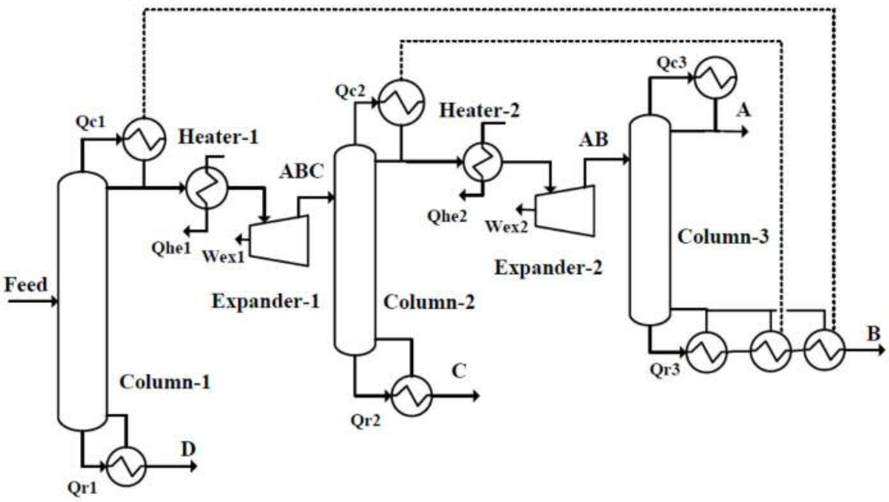


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Fig. 4. Indirect sequences: A is Indirect(1X2&2X3), B is Indirect(C1&C2-R3)(1X2&2X3)

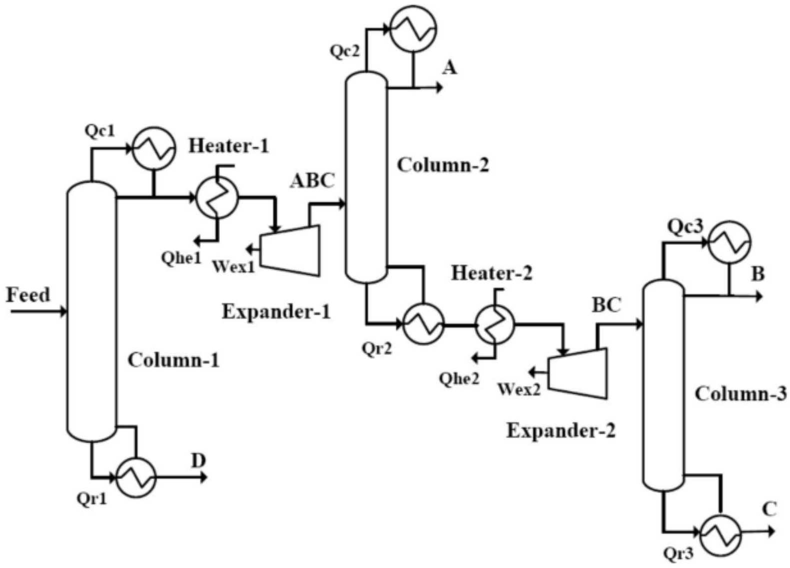

Fig. 5. Indirect-Direct(1X2&2X3) sequence

The possible sequences for this process are as follows: Distributed sequence (Figure 1): AB/CD; A/B; C/D. Direct sequence (Figure 2): A/BCD; B/CD; C/D. Direct-indirect sequence (Figure 3): A/BCD; BC/D; B/C. Indirect sequence (Figure 4): ABC/D; AB/C; A/B. Indirect-direct sequence (Figure 5): ABC/D; BC/D; B/C. The novelty of this work is to find the best sequence with its heat integrations and power productions in order to achieve the minimum value of total annual cost.

2. Methodology

The target of this research is to find the best heat integrated sequence with maximum potential of power generation to reach the minimum of total annual cost for the considered industrial case study. To reach this aim, two integration steps are proposed. In the first step, all the simple sequences and heat integrated sequences are designed and optimized. In the second one, the ability of power generation in each sequence is considered and applied.

2.1 Step 1: heat integration

In the first step, the target is to minimize the total annual cost of sequences by creating heat integration among columns. To calculate the temperatures and heat flow rates of condensers and reboilers and find the design specifications of columns such as the actual number of trays, reflux ratio, rectify vapor flow, and etc., all the five simple sequences are simulated by shortcut and rigorous method of commercial software [AspenHYSYS ver.2006]. NRTL equation of state was used for simulation in all cases. The base case was compared to industrial case study, Maroon Petrochemical plant, and other sequences were done by optimization in the base case. Then, the equations of diameter and height of columns, annual capital cost of columns and heat exchangers and annual operation cost are modeled and used to calculate the total annual cost of each sequence. The equations used to account diameter and height of columns are (Couper et al. 2005):

$$D = 5e-3 \times \sqrt[0.25]{MWHC \times (1.8 \times TBP + 492)} \times \sqrt[0.5]{RV} \quad (2)$$

$$H = (Nt-1) \times 0.6 + 6 \quad (3)$$

The height of column (Equation 3) is evaluated by considering 0.6 m as tray spacing and 6.0 m as disengagement.

The basic equations for calculating the capital cost are the total cost of distillation columns and heat exchangers (Couper et al. 2005). The total cost of a distillation column can be considered as a sum of the costs of column shell and trays. The equation for calculating the column shell is:

$$\$ = \left(\frac{M\&S}{280}\right)(937.61)D^{1.066}H^{0.802} \quad (4)$$

All the correlations are valid for carbon steel construction and are updated from mid-1968 to 2011 by utilizing the Marshall and Swift cost index (Chemical Engineering Magazine). The correlation (E.4) is valid for a pressure less than 345 kPa, otherwise a correction factor $\left[1 + 1.45 \times 10^{-4}(P-345)\right]$ must be applied.

For calculating the cost of column's trays, considering sieve trays, the following correlation has been utilized:

$$\$ = \left(\frac{M\&S}{280}\right)(97.24)D\ H^{0.802} \quad (5)$$

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

$$A = \frac{Q}{U\ \Delta T_{LMTD}} \quad (6)$$

Mean values of 1800 W/ (m^2 K) and 2100 W/ (m^2 K) 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:

$$\$ = \left(\frac{M\&S}{280}\right)(474.67)A^{0.65} \quad (7)$$

The reported correlation is valid for an exchange area range between $18.6 < A < 464.5\ m^2$.

The capital cost is annualized over a period which is often referred to plant life time. In this study, the plant lifetime is 10 years.

Annual operating cost is calculated by the amount of heat flow rates of condensers and reboilers of columns and utility costs of Table 1. Heat flow rates of condensers and reboilers are calculated by simulation. The time fraction of operation for annualizing the operating cost is 8 000 h/a.

In next step, after calculating the TAC function, the optimizer toolbar of commercial software is used to minimize the objective function of TAC according to the continue parameters of column's pressures and the temperature constraints of process. The optimization method SQP of the optimizer toolbar is used for minimizing the problem. With the continuous decision for the individual columns' pressure {p}, the objective function formulation is:

$$\text{Min TAC} = \min C(\{p\}),\ \{p\}\ \varepsilon\ P \tag{8}$$

where C( ) represents the total annual cost model for the system, P is the vector of operating pressures for all individual columns in the system, and P can be a set of feasible range of {p}. For minimizing the TAC function of each simple sequence, continues parameters of column's pressures can change in the range of 1 bar to 30 bars. For temperature constraints, temperatures of columns have to be between the temperatures of cooling water and the temperature of hp-steam.

After the optimization of the simple sequences, in the next step, the possible heat integration in each sequence is found. In each sequence, there are 3 condensers (C1, C2, C3) and 3 reboilers (R1, R2, R3) which can transfer heat to the other according to their temperatures. For each sequence, there would be 12 different situations for heat integration:

C1→R2, C1→R3, C1→R2&R3, C2→R1, C2→R3, C2→R1&R3, C3→R1, C3→R2, C3→R1&R2, C1&C2→R3, C2&C3→R1, C1&C3→R2

For example, the situation C1→R2 means that the condenser C1 transfers heat to reboiler R2, according to the minimum temperature difference $\Delta T_{min}=10$ K. To find that the condenser C1 can transfer heat to reboiler R2, we increased the pressure of column1 to the allowable maximum pressure point, according to the temperature constraint, to increase the temperature of the condenser C1 to its maximum level. On the other side, we decreased the pressure of column 2 to allowable minimum pressure, according to the temperature constraint, to decrease the temperature of reboiler R2. If the temperature of condenser C1 was higher than the temperature of reboiler R2 plus ΔT_{min}, the heat integration of C1→R2 could be possible. Now for this situation, a new temperature constraint ($T_{C1}-T_{R2}>10$) and some changes to the calculation of total annual capital cost and annual operation cost are added to the optimization procedure of the objective function of TAC in optimizer toolbar. The optimum pressures of columns and the minimum TAC are obtained by the optimization, and the data and structure of sequence with heat integration of C1→R2 are stored. Totally, for all five simple sequences, there would be 60 different situations for heat integration.

This procedure is done for all the mentioned different situations of heat integration for each simple sequence, and all the possible heat integrated sequences are optimized. Finally, the optimum simple sequences and heat integrated sequences are compared to find the best structure with the minimum total annual cost.

2.2 Step 2: power generation

In the second step, the target is to find the possible power generation in all the sequences which were designed and optimized in step 1. This investigation is in theory and is just a feasibility study to find the effects of power generation on the exergetic efficiency of process and total annual cost.

First, in each sequence, the streams between columns with negative pressure difference ($\Delta P < 0$) must be found. It means if there is a process stream between column 1 and column 2 and the pressure of column 1 is higher than column 2, this stream has $\Delta P < 0$ and the potential of power generation by installing expander. We tried to increase the ΔP between two columns according to the feasible pressure boundaries, and created heat integration among columns to have a rational pressure difference for power generation. In each sequence, all suitable streams are found and expanders are installed on the streams. The Expander capital cost evaluation is based on the power generation evaluated by utilizing the usual design formula [20]:

$$\$ = \left(\frac{M \& S}{280}\right)(0.298)\left(\dot{W}\right)^{0.81}(1000) \tag{9}$$

The reported correlation is valid for power generation between 26.8 kW $< \dot{W} <$ 6705 kW.

The capital cost of expander is modeled and added to the total annual cost function. To increase the power generation of expander, the intake stream is superheated by hot utilities. The capital cost of the heat exchanger is also modeled and added to the TAC. The cost of consumed hot utility for superheating and the benefit of power generation are calculated and inserted to the total annual cost. After installing these equipments and inserting their capital and operating cost to the objective function of TAC, in the next stage, the TAC function is minimized by the optimizer toolbar of commercial software. The continuous parameters for the optimization are the heat flow rates which are used for superheating the inlet streams to expanders. The outlets of expanders have to be saturated vapor; this constraint should also be considered in the objective function optimization. By minimizing the objective function and finding the optimum parameters, new sequences with power generation between columns are designed. This procedure is done for all the optimum sequences obtained from step 1.

At the end of two Steps, all the designed complex sequences with heat integration and power generations are compared to find the best separation structure with the minimum total annual cost.

For each sequence, the exergetic efficiency is calculated to consider the effects of power generation on the efficiency of the separation sequences (Kotas 1995). Kotas equation is used to calculate the exergetic efficiency (η_{ex}):

$$\eta_{ex} = \frac{\sum \Delta E_{out}}{\sum \Delta E_{in}} \tag{10}$$

This equation is based on the ratio of the required output exergy (ΔEout) of the process to the input exergy (ΔEin) necessary to obtain that output. The exergies of flows have also been obtained by using the following relationship:

$$E_X = E_{ph} + E_{ch} \tag{11}$$

where E_{ph} and E_{ch} are the physical exergy and the chemical exergy of a stream respectively. Ambient temperature and pressure are 30 ^{o}C and 1 bar.

Table 1. Utility costs

Utility	Temperature Level (°C)	Values
LP-steam ($/t)	158	13
MP-steam ($/t)	200	16
HP-steam ($/t)	250	20
Cooling water ($/t)	35-45	0.082
Electricity ($/kW h)	-	0.1

Table 2. Feed and product specifications

Product	Feed		
Product groups	Components		Xi (Mol Fraction)
A	Ethane	C2H6	0.0001
	M-Acetylene	C3H4	0.0047
	Propene	C3H6	0.4538
B	Propane	C3H8	0.2912
	VnylAcetlen	C4H4	0.0003
C	DMAcetylene	C4H6	0.0011
	i-Butene	C4H8	0.1541
	n-Butane	C4H10	0.0359
D	n-Pentane	C5H12	0.0419
	n-Hexane	C6H14	0.0137
	n-Heptane	C7H16	0.0031

Table 3. Optimal schemes

Description	Primary Distributed			Distributed(1X2&1X3)			Distributed(C1&C3-R2) (1X2&1X3)		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3	Col. 1	Col. 2	Col. 3
Pressure (kPa)	1785	1940	735	2998	2139.43	650	2989	2217.96	1042
Diameter (m)	3.35	4.89	1.67	3.33	5.04	1.86	3.36	5.04	1.89
Reflux ratio	0.71	8.52	0.64	0.69	9.09	1.02	0.73	9.12	1.1
Actual plates	54	163	43	45	188	40	43	187	42
Total actual plates		260			273			272	
Heating rate (kJ/h)		112045868.6			145598914			111260285.8	
Cooling rate (kJ/h)		118136557.5			121775816			87013058.14	
HX heat flow rate (kJ/h)		-			-			34026538.26	
Power generate (kW)		-			6604.35			6321.32	
Steam cost ($/a)		4560843.99			6962706.94			5659338.3	
C.W cost ($/a)		1854009.13			1911122.85			1365563.78	
Power cost ($/a)		-			5283477.77			5057053.02	
Operating cost ($/a)		6414853.12			3590352.03			1967849.07	
Capital cost ($/a)		762531.12			881240.89			865687.11	
TAC ($/a)		7177384.24			4471592.92			2833536.17	
Operating cost saving (%)		-			44.03			69.32	
Capital saving (%)		-			-15.57			-13.53	
TAC saving (%)		-			37.7			60.52	
TAC saving by heat (%)		-			-			27.86	
TAC saving by power (%)		-			37.7			32.66	
Exergetic efficiency		0.28			0.97			0.89	

Description	Direct(1X2&2X3)			Direct(C3-R1)(1X2&2X3)			Direct(C3-R2&R1)(1X2)		
	Col.1	Col. 2	Col.3	Col.1	Col. 2	Col.3	Col. 1	Col. 2	Col. 3
Pressure (kPa)	2979	2413	651	2414	1968.52	1638	2564	1838.52	2289
Diameter (m)	6.19	3.06	1.85	6.23	3.02	1.92	6.25	3.01	1.72
Reflux ratio	8.58	2.64	1.03	8.77	2.52	1.19	8.82	2.49	0.76
Actual plates	180	44	40	174	43	46	177	43	50
Total actual plates		264			263			270	
Heating rate (kJ/h)		129162671.4			122821995			118241776.6	
Cooling rate (kJ/h)		110008212.6			102951263			102380517.2	
HX heat flow rate (kJ/h)		-			9894895.56			7414038.13	
Power generate (kW)		4900.3			4676.78			3339.29	
Steam cost ($/a)		6407800.18			6117755.63			5747236.03	
C.W cost ($/a)		1726444.68			1615694.46			1606737.3	
Power cost ($/a)		3920240.15			3741423.2			2671432.86	
Operating cost ($/a)		4214004.71			3992026.89			4682540.48	
Capital cost ($a)		798318.19			812172.66			769527	
TAC ($/a)		5012322.9			4804199.55			5452067.48	
Operating cost saving (%)		34.31			37.77			27	
Capital saving (%)		-4.69			-6.51			-0.92	
TAC saving (%)		30.17			33.06			24.04	
TAC saving by heat (%)		9.85			13.14			11.1	
TAC saving by power (%)		20.32			19.92			12.94	
Exergetic efficiency		0.62			0.59			0.4	

Table 3. Optimal schemes (continue)

Description	Direct-Indirect(1X2&2X3)			Direct-Indirect(C2-R1)(2X3)			Indirect(1X2&2X3)		
	Col. 1	Col. 2	Col. 3	Col. 1	Col. 2	Col. 3	Col. 1	Col. 2	Col. 3
Pressure (kPa)	2550.8	2115	1804	2024	2740	1717	2708	2407.57	2077.7
Diameter (m)	6.24	2.63	2.24	6.2	2.42	2.23	3.17	0.74	5.03
Reflux ratio	8.81	0.62	1.97	8.68	0.37	1.94	78.96	1.11	9.07
Actual plates	177	45	42	174	49	42	54	48	192
Total actual plates		264			265			294	
Heating rate (kJ/h)		143767876.8			116638106			182277792.9	
Cooling rate (kJ/h)		117168747.3			102358285			141432413.4	
HX heat flow rate (kJ/h)		-			14684071.1			-	
Power generate (kW)		6412.31			3242.08			10233.16	
Steam cost ($/a)		7183378.15			5746841.97			9252759.21	
C.W cost ($/a)		1838820.53			1606388.39			2219609.17	
Power cost ($/a)		5129847.45			2593667.56			8186528.94	
Operating cost ($/a)		3892351.23			4759562.81			3285839.43	
Capital cost ($/a)		880516.89			864274.22			1076985.23	
TAC ($/a)		4772868.12			5623837.03			4362824.66	
Operating cost saving (%)		39.32			25.8			48.78	
Capital saving (%)		-15.47			-13.34			-41.24	
TAC saving (%)		33.5			21.65			39.21	
TAC saving by heat (%)		1.2			6.48			-17.31	
TAC saving by power (%)		32.3			15.17			56.52	
Exergetic efficiency		0.73			0.43			0.9	

Description	Indirect (C1&C2-R3)(1X2&2X3)			Indirect-Direct(1X2&2X3)		
	Col. 1	Col. 2	Col. 3	Col. 1	Col. 2	Col. 3
Pressure (kPa)	2999	2606	2074	2710	2373	1812
Diameter (m)	3.2	3.02	5.04	3.18	5.24	2.24
Reflux ratio	0.24	1.14	9.11	0.23	9.59	1.98
Actual plates	59	56	188	52	198	42
Total actual plates		303			292	
Heating rate (kJ/h)		128594616.2			168527055	
Cooling rate (kJ/h)		88189888.36			133309415	
HX heat flow rate (kJ/h)		53603010.61			-	
Power generate (kW)		9944.24			8890.03	
Steam cost ($/a)		7572733.65			8691617.51	
C.W cost ($/a)		1384032.7			2092128.61	
Power cost ($/a)		7955389.1			7112026.26	
Operating cost ($/a)		1001377.25			3671719.87	
Capital cost ($/a)		1189255.27			943752.72	
TAC ($/a)		2190632.52			4615472.59	
Operating cost saving (%)		84.39			42.76	
Capital saving (%)		-55.96			-23.77	
TAC saving (%)		69.48			35.69	
TAC saving by heat (%)		17.15			-8.11	
TAC saving by power (%)		52.33			43.8	
Exergetic efficiency		0.85			0.8	

3. Case study

The industrial process examined in this paper is the high temperature separation section of olefin plant which separates a multi-component flow to four categories of products with a purity of 0.98. The intake flow has a temperature of 59.39 ^{o}C and a pressure of 1899.5 kPa, and enters the process with an amount flow rate of 1355 kmol/h. The compositions of the process intake flow and groups of products are shown in Table 2. The sequence of the initial process was the distributed type and is shown in Figure 1-A, and is regarded as a basis for comparison with other types. A number of different complex sequences obtained with their heat integration and power generation are shown in Figures 1 to 5. The results of design and optimization are shown in Table 3.

4. Discussion and results

The distributed sequence of the initial process is chosen as the base for comparison of other sequences (Figure 1-A). The exergetic efficiency of the initial process is 0.28 (Table 3).

From Table 3, it can be seen that the sequence Indirect (C1&C2-R3)(1X2&2X3) (Figure 4-B) presenting a TAC optimization of 69.48% provides the best result when compared to other sequences. In this sequence, two expanders have been installed between the columns and the intake flows of expanders have been superheated by using the HP steam (Figure 4-B). The result of the TAC function showed that using HP steam for heating the intake flows of the expanders in sequences is preferred to LP and MP steams because of high power generation resulting from expanders and utility cost factors. Two heat integrations are carried out between the condensers of columns 1 and 2 and the reboiler of column 3. In this case, operational costs have been reduced by 84.39 %; capital costs have been augmented by 55.96 %, and TAC has been lowered by 69.84 % (Table 3). The improvement caused by the heat integration was 17.15 %, and the improvement resulted from power production was 52.33 %. Since we have two heat integrations in the processes, and thus, high power production in expanders, the operational costs have had a considerable reduction. The power production is so high in this case because there is high flow rates going through the expanders put between the columns. In this case, separation of products A and B in column 3 has required a high flow rate for producing power between the columns since A and B form 45 % and 29 % of the total flow rate of the process, respectively. However, the fact that these products exist with high flow rates in columns has greatly increased the heat consumption of the process. The exergetic efficiency is 0.85 in this case which is a result of high ratio of power production to heat consumption and low temperature of heat transfer in the process.

The second best sequence is Distributed (C1&C3-R2) (1X2&1&3) with TAC optimization of 60.52 %. In this sequence, power is produced by putting two expanders in the way of AB and CD flows (Figure 1-C). Saturated liquid streams of column 1 have been superheated by using HP steam. In this case, two heat integrations have been done between condensers of columns 1 and 3 and the reboiler of column 2. Compared to the base case which is a result of the optimization caused by two heat integrations and two expanders installed in the process (Table 3), the operational cost has been improved by 69.32 %. The exergetic efficiency in this sequence is 89%. Condenser heats of columns 2 and 3 have increased and their reboiler heats have decreased due to steam flow intake to these columns, and since the condenser heat of columns 1 and 3 have been transferred to the reboiler of column 2, the heat consumption of the reboiler of column 2 has greatly decreased. In fact, the preheating of the heat flow to column 3 has increased the heat transfer from this column to column 2. Reflux ratio and diameters of columns 2 and 3 have also increased when compared to the state without expander due to the increase of steam volume. The initial cost has shown a 13.53 % increase when compared to the base state which is chiefly because of costs of the employed expanders. The number of trays of column 2 has also increased due to steam volume and reflux ratio of the column since when the steam volume increases; one has to increase the number of trays to enlarge the contact area of vapor and liquid. Generally TAC has been improved by 60.52 %. It involves a thermal improvement of 27.86 % and the optimization caused by power production of 32.66 %. In this case, the expanders have also produced a high power due to high pressure difference and high flow rate between the columns and superheating of the intake flow of expanders.

The third best sequence with TAC optimization of 56.52 %. is Indirect(1X2&2X3) which has two expanders between the columns (Figure 4-A). Operational costs have decreased by 48.47 %. Capital costs increased by 41.24 % and TAC reduced by 39.21 % (Table 3). 17.31 % of the increase in TAC was caused by changing the sequence and 56.52 % of the decrease of it was caused by the power produced by expanders. The exergetic efficiency of this case is 90% representing a very high power production in the process. In this case, separation of products A and B in column 3 has required a high flow rate for producing power between the columns since A and B form 45 % and 29 % of the total flow rate of the process, respectively. However, the fact that these products exist with high flow rates in columns has greatly increased the heat consumption of the process. In general, TAC has largely reduced because of high power production in the process which shows the importance of the profits brought about by power production when compared to the costs imposed by more heat consumption.

Separation of products A and B, which have high flow rates and also approximately the same relative volatilities, in column 3 has increased the possibility of heat integration in sequence Indirect (C1&C2-R3)(1X2&2X3) (Figure 4-B), because of low temperature difference in column 3. Existence of these components in the columns 1 and 2 has also caused the lower temperature difference in other columns, which increases the possibility of heat integration in sequence. On the other hand, since these two products include almost 75 % of the total input flow rate of the process, the power production between the columns has also been high when compared to other states. A negative point of this sequence which is an example of consecutive sequences is the high volume of steam in columns. This point increases the costs related to heat consumption of columns when compared to Distributed cases. Placing expanders between the columns not only produces power, but also it brings about the saturated vapor into the columns and thus reduces heat consumption of reboilers. One can say by creating suitable conditions in the process for power production, we can greatly influence operational costs.

In the distributed sequences, the pressure difference between the two columns like 1 and 2 has no effect on the pressure difference between the columns 1 and 3, that is one can increase the pressure of column 1 and decrease the pressures of columns 2 and 3 to create a high pressure difference between the columns. It shows the high freedom of these sequences for creation higher pressure differences between columns. The fact that the columns are sequential in direct or indirect cases has resulted in the pressure difference between two columns influencing the pressure difference of the two other columns. For instance, in indirect sequences, by decreasing the pressure of column 2 in order to increase the pressure difference of the two subsequent columns 1 and 2, the pressure difference between columns 2 and 3 is reduced if assuming that column 1 has the highest pressure and column 3 has the lowest pressure. It decreases the freedom of these sequences for creating higher pressure differences between columns.

A high temperature difference inside the columns is one of the factors making changing their pressures difficult. The reason for that is to have materials with largely different volatilities in the column. Therefore, in order to have more freedom in changing pressures of columns and creating heat integration, we should select sequences in which there is a low temperature difference between condensers and reboilers of the columns.

High flow rate of flows between the columns increases the process's power production. However, it also increases the volume of steam inside the columns and thus their heat consumption. In the past, lots of works tried to separate the components with high flow rates in primary distillation columns, because separation of these components at the end of sequence increases the heat consumption of the primary and intermediate columns. Since producing power inside the process has a high profit compared to steam costs, noticing power production is of considerable importance in designing sequences. Therefore, the best sequence should not only convey a low heat consumption and high freedom in creating heat integration, but also it should include high flow rate flows between the columns for increasing power generation.

5. Conclusion

In this study, all possible sequences with heat integration and power generation for separation of a multicomponent flow to four categories of products are considered by the two presented integration steps. The results represent a very high improvement for the sequence Indirect (C1&C2-R3) (1X2&2X3) since the properties of the intake flow to the process are in a way that in this sequence, not only do we have a high freedom for carrying out heat integration, but also a large amount of power is produced between the columns because of having high flow rate flows between the columns. The best sequence should not only have a low heat consumption and high freedom for creating heat integration, but also it should include high flow rate flows between the columns for increasing the potential of power generation which can highly increase the exergetic efficiency of the sequence and decrease the total annual cost of the process. Regarding the calculated results, the following two rules of thumb should be considered while designing a sequence:

- Choose sequences with minimum sums of temperature differences of columns
- Choose sequences with maximum sums of flow rates between the columns

Nomenclature

Q	*Heat flow rate*
p	*Pressure*
TBP	*Boiling Point Temperature*
ΔT_{min}	*Minimum temperature difference*
ΔT_{LMTD}	*Logarithmic Minimum Temperature Difference*
ΔP	*Pressure difference*
$\dot{W}$	*Power generation*
RV	*Rectify vapor rate*
Xi	*Mole Fraction*
$MWHC$	*Molecular Weight of Heaviest Component*
Nt	*Number of trays*
D	*Diameter of column*
H	*Height of column*
A	*Area of heat transfer in heat exchanger*
U	*Heat transfer coefficient of heat exchanger*
S_n	*Number of simple sequence*
n	*Number of components*
$\{p\}$	*Decision pressure parameter*
TAC	*Total Annual Cost*
E_X	*Exergy of flow*
E_{ph}	*Physical exergy*
E_{ch}	*Chemical exergy*
E_X	*Exergy of flow*
E_{ph}	*Physical exergy*
E_{ch}	*Chemical exergy*
ΔE_{in}	*Input exergy*
ΔE_{out}	*Output exergy*
η_{ex}	*Exergetic efficiency*
K	*Kelvin*
$R1$	*Reboiler of column 1*
$R2$	*Reboiler of column 2*
$R3$	*Reboiler of column 3*
$C1$	*Condenser of column 1*
$C2$	*Condenser of column 2*
$C3$	*Condenser of column 3*
HX	*Heat Exchanger*
SQP	*Successive Quadratic Programming*
CW	*Cooling Water*
Min	*Minimum*
$M\&S$	*Marshall and Swift Cost Index*
E_X	*Exergy of flow*
E_{ph}	*Physical exergy*
E_{ch}	*Chemical exergy*
ΔE_{in}	*Input exergy*

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