The effect of middle components purity in quaternary mixture separation using distillation system

Jamal Ahmed Enghita
Department of Chemical Engineering, Faculty of Engineering (Algaraboli), Elmergib University
Corresponding Author Email: enghitaj@gmail.com

Abstract

Separation processes play critical roles in industry, including the removal of impurities from raw materials and the overall separation processes accounts for 40-70% of both capital and operating costs in industry. The main objective of this study was exploration of design and economic features of conventional distillation schemes for separation of quaternary mixture (Propane, i-Butane, n-Butane, i- Pentane) with high products separation of 95%, 90%, 90%, and 95%, respectively for the equimolar feed composition. The studied distillation schemes are: the effect of middle components purity (i-C4 and n- C4) is also on energy and capital investments, conventional distillation scheme, The middle components purity was changed from 90% to85%and 80 %.The results of economic study show conventional Distillation cases studied at different composition of middle components purity at (0.9, 0.85and 0.8) and we get highest saving on the case of 80% side stream composition with 19 %saving in energy and 17%saving in TAC, and the saving on case of 85% side stream composition with 12% saving in energy and 11 % saving in TAC, compared with case of 90%middle component purity.

Keywords: Separation processes, Conventional distillation column, Process optimization, Quaternary mixture, Separation.

Introduction

There are many separation technologies such as distillation, extraction, adsorption, crystallization, and membrane-based technologies. However, many considered the distillation technique is a mature technology and represents 70% of all separation systems [1]. In addition, distillation is the largest single energy consumer in the chemical process industries but distillation does not consume energy but degrades the heat input to the reboilers that is subsequently rejected in the condenser. The most effective way to reduce the energy consumption of distillation is by effective heat integration. Design and operation of the distillation are to be considered simultaneously with its heat integration. For single distillation columns, it is straightforward to identify appropriate heat integration opportunities. For complex distillation systems, the most appropriate combination of distillation system design, operation, and heat integration are far from straightforward. The whole separation system together with its heat integration and utility system must be considered simultaneously. The rapid increase of energy prices in the past decades has given motivation to many efforts to conserve energy. To conserve energy, the ideas of component purity and heat integration have received more attention.

Introducing thermal couplings to eliminate intermediate reboilers and condensers to save energy and capital cost, improving operability of thermally coupled columns, enabling double and multi-effect distillation of thermally coupled configurations, performing simultaneous heat and mass integration, and conducting any thermally coupled distillation in n-product streams are very important issues in separation processes. Many workers have studied distillation configurations for multi-component distillation systems [2]. Thermal couplings configurations have been known for a while, but it was not used until the 1980s particularly when configurations started to receive more attention due to their potential for large heat duty savings and cost effectiveness. Improving energy efficiency in refineries and Gas plant has been studied extensively in the realm of production management due to the potential of large energy savings and gas emission reduction. Sittig has reported energy saving for petroleum refinery industry and suggested the use of intermediate reboilers in crude towers can reduce the heat load on the furnace in order to increase the energy efficiency of the system [3]. Similar work is conducted to study and concluded that the installation of preflash units or prefractionator columns to existing crude oil distillation installations could save energy and allow the furnace to reduce its utility consumptions [3] In 1994, Dhole and Buckingham have published a methodology based on pinch analysis for saving energy in distillation units [4]. Many approaches have been applied to optimize CDUs with the consideration of economic efficiency and environmental impact They are mainly divided into two categories. One is based on the heat integration analysis to reduce energy consumption by retrofitting heat exchanger networks [4] . They reported results showed an increase in heat recovery of the heat exchanger network to reduce the utility consumption in the process furnaces. Most of the work dealed with simple configuration(sequences of conventional one-feed, two product columns).The main objective of the present study was exploration of design and economic features of conventional distillation schemes for separation of quaternary mixture of Propane, i-Butane, n-Butane, and i- Pentane with high products separation of 95% propane and 95% of i-pentane. The present work is based on studying and rigorous modeling (design and simulation) of some conventional schemes for the separation of quaternary mixture of light hydrocarbons(C3, i-C4, n-C4 and i-C5). Economic evaluation of the distillation schemes is also studied.

Design of Distillation Systems

Different types of distillation systems are studied and compared to the conventional direct scheme. The Conventional Direct Sequence given by Elaahi and Luyben [12] is designed, simulated, and economically evaluated in term of energy saving and TAC. In case of quaternary mixture separation, the designed distillation systems consisted of three distillation columns and direct separation sequence is analyzed. Each column has feed stream, top product (distillate), and bottom product, in addition to total condenser and reboiler for each one. The studied distillation systems are selected from different types of middle component purity (90%, 85%,% and 80%). The sequence is direct in which the lightest component is taken overhead in each column, the bottom of the first column is fed to the second column, and the bottom of the second is fed to the third column. Figure 1 shows the designed conventional direct distillation sequence with total condensers. The conventional scheme is considered as the base case scheme for all the investigated schemes to compare the percentage of savings in both energy and total annual cost (TAC).

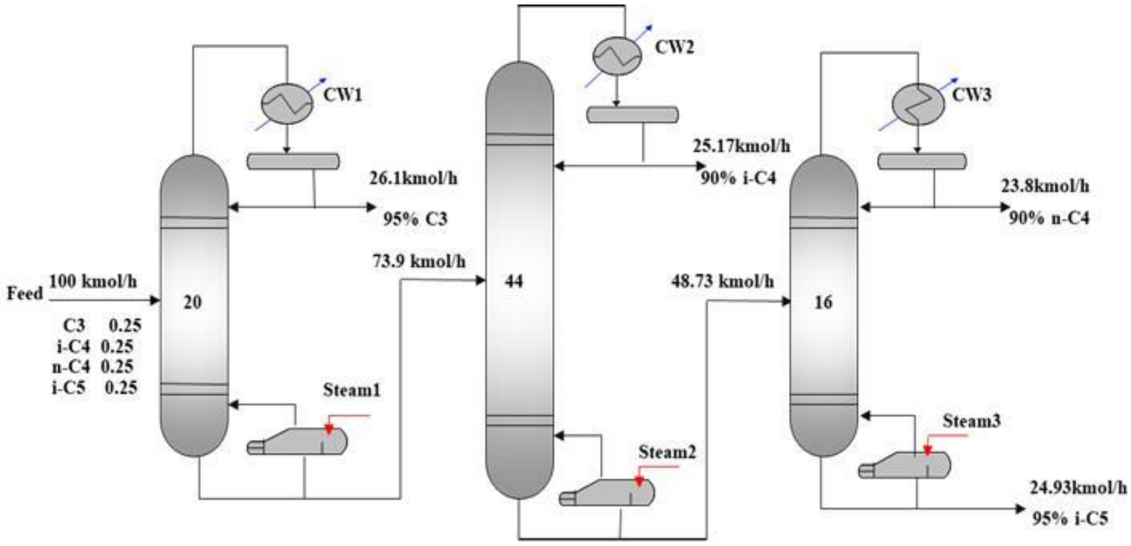

Figure 1: Conventional direct distillation sequence with total condensers.

Cases Study

The separation of quaternary mixture consisted of Propane, i-Butane, n-Butane, and i-Pentane by several distillation schemes for 95 mol% product purity for both of Propane and i-Pentane in addition, of 90% for i-Butane and n-Butane. The feed composition consisted of same mole percentage of the studied hydrocarbons (25%) for case 1. The feed composition was saturated liquid hydrocarbons at 50 °C with flow rate of 100 kmol/h. The feed and product specifications of the studied rigorous schemes are given in Table 1.

Table 1: Feed and product specifications for the case study

Components	Feed stream	Product streams			
		A	B	C	D
	fraction	fraction	fraction	fraction	fraction
Propane	0.25	0.95	0.09	0.00	0.00

i-Butane	0.25	0.04	0.90	0.09	0.00
n-Butane	0.25	0.01	0.01	0.90	0.05
i-Pentane)	0.25	0.00	0.00	0.01	0.95
Total	1.00	1.00	1.00	1.00	1.00

For a certain number of trays, the reflux ratio is calculated by the HYSYS. The feed location to each column is adjusted by changing the location up and down until the lowest value of the reflux ratio is achieved at the specified product purities.

Estimation of trays Number, Reflux Ratios and Feed Location

1- Fenske equation: For multi component systems, an approximate value of the minimum number of equilibrium stages (at total reflux) may be obtained from the Fenske equation. In the Fenske equation, given below as Equation 1, relative volatility is based on the light key relative to the heavy key[9].

$$N_{\min} = \frac{\ln\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D \left(\frac{x_{HK}}{x_{LK}}\right)_B\right]}{\ln\left(\alpha_{LK/HK}\right)_{av}}$$

[1]

Where $N_{\min}$ = minimum number of equilibrium stages

x_{LK} = mole fraction of light key

x_{HK} = mole fraction of heavy key

D = when used as subscript, denotes distillate product

B = when used as subscript, denotes bottom product

$(\alpha_{LK/HK})$= average value of relative volatility of light key relative to heavy key

The average value of relative volatility is calculated from the mean value of relative volatility at the top of the column $(\alpha_{LK/HK})_D$ and at the bottom of the column $(\alpha_{LK/HK})_B$. $(\alpha_{LK/HK})_D$ is based on the dew point temperature of the overhead vapor whereas $(\alpha_{LK/HK})_B$ is based on the bubble point temperature of the bottom liquid.

$$(\alpha_{LK/HK})_{av} = \sqrt{(\alpha_{LK/HK})_D (\alpha_{LK/HK})_B} \qquad [2]$$

Since the distributions of the non-key components in the distillate and bottoms are not known, determinations of the dew and bubble points are partially trial and error.

2- Minimum reflux by Underwood method: For multi-component mixtures, the Underwood method may be used for estimating minimum reflux ratio. Underwood proposed Equations (3) and (4) for calculation of minimum reflux ratio[9]:

$$\sum_{i=1}^{n} \frac{\alpha_i x_{f_i}}{\alpha_i - \theta} = 1 - q \qquad [3]$$

Where n = number of components

α_i = average relative volatility based on component i relative to heavy key

x_{f_i} = mole fraction of component i in feed

q = number of moles of saturated liquid produced on the feed tray per mole of feed

θ = unknown parameter to be determined by trial and error

The correct value of θ will be between relative volatility of the two key components. After the value of θ is determined, Equation (4) may be used to determine minimum reflux ratio.

$$R_{\min} + 1 = \sum_{i=1}^{n} \frac{\alpha_i x_{D_i}}{\alpha_i - \theta} \qquad [4]$$

Where $R_{\min}$ = minimum reflux ratio ,

x_{D_i} = mole fraction of component i in distillate (D)

3- Number of equilibrium stages by Gilliland correlation: Once minimum stages and reflux ratio are known, the number of equilibrium stages may be determined as a function of selected values of operating reflux ratio. Several methods have been proposed including the correlation of Gilliland to determine the number of equilibrium stages as a function of reflux ratio.

The Gilliland correlation was first developed as a plot of $\frac{N - N_{\min}}{N+1}$ versus , but latter was transformed by (Eduljee, 1975) into Equation [9]

$$\frac{N - N_{\min}}{N+1} = 0.75\left(1 - \left[\frac{R - R_{\min}}{R+1}\right]^{0.566}\right) \qquad [5]$$

Where N = number of equilibrium stages , $N_{\min}$ = minimum number of equilibrium stages

R = operating reflux ratio , $R_{\min}$ = minimum reflux ratio

Results of the Shortcut Analysis

For a given feed and product specifications (Table 2) and after material balance around the studied sequence, the unknown composition of each component is determined, and then the light and heavy keys can be specified. After that, the minimum number of trays (Fenske equation), minimum reflux ratios (Underwood equation), and the feed stage locations (Kirkbride equation) are estimated. The results of shortcut method for case 1 are summarized and given in Table 2 .

Table 2: Shortcut method results of case 1

Parameter	Column 1	Column 2	Column 3
$N_{\min}$	10	22	8
$R_{\min}$	3.72	6.31	2.25
$R = 1.1\, R_{\min}$	4.09	6.941	2.475
$N = 2\, N_{\min}$	20	44	16
M	10	22	8
P	10	22	8

Results of Rigorous Simulation Schemes

According to the TAC calculation in the present case study, the plant life operating time was selected to be 10 years with 8000 operating hours per year.

Case 1: The results of Base case (Conventional direct distillation sequence with 90% middle component purity) are given in Table 3. In this separation process, It notice a relative volatility (αav) of the columns αav3 > αav1 > αav2. This means that the trays of column 2 to separate the products (B,C, and D) need more than other columns and also the column 2 consume high energy compared to the other columns. The column 2 consumed high energy and increase the highest cost in both, capital and operating cost with respect to the other two columns.

Table 3: The base case

	Column 1	Column 2	Column 3
Column pressure (kpa)	*1455*	652.4	456.9
Column diameter (m)	0.84	0.884	0.5455
Aave	1.781915	1.292199	2.088117
Reflux ratio	6.69	8.3	2.52
Condenser duty (KJ/hr)	3.79E+06	4.40E+06	1.68E+06
Reboiler duty (KJ/hr)	2.53E+06	3.91E+06	1.58E+06
Theoretical number of stages	*20*	40	14
Overall column efficiency	82	76	70

Actual number of stages	*26*	60	24
Annual capital cost ($/yr)	2.98E+04	4.11E+04	1.70E+04
Total annual capital cost ($/yr)		8.79E+04	
Cooling water cost ($/yr)	1.28E+04	1.48E+04	5.64E+03
steam cost ($/yr)	1.73E+05	2.68E+05	1.08E+05
Operating cost ($/yr)	1.86E+05	2.82E+05	1.14E+05
Total operating cost ($/yr)		5.82E+05	
TAC ($/yr)	2.16E+05	3.23E+05	1.31E+05
TAC of the whole sequence ($/yr)		6.70E+05	
Total annual cost saving %		0	
Total operating cost saving %		0	
Total QR (KJ/hr)		8.02E+06	
Total QR saving %		0	
TAC saving %		0	

Case 2: From Table 5 shows the effects of middle component purity (MCP), i-C4 and n-C4, of the composition of 0.85 for the previous scheme on energy consumption. , Distillation Sequence with 85% MCP where the reboiler duty of the second and third column shows the reduction in energy consumption from 3.91E+06 kJ/h to 2.98+06kJ/h and from 1.58E+06 kJ/h to 1.51E+6 kJ/h, respectively. These effects are translated into energy saving of 12.4 % and TAC saving of 11.45 % as shown in Table 4.

Table 4: Middle component purity for Case 2 (0.85).

	Column 1	Column 2	Column 3
Column pressure (kpa)	*1455*	654.9	456.9
Column diameter (m)	0.84	0.8	0.542
αave	1.960013	1.355843	2.385092
Reflux ratio	6.3	7.20	2.1
Condenser duty (KJ/hr)	3.80E+06	3.47E+06	1.61E+06
Reboiler duty (KJ/hr)	2.53E+06	2.98E+06	1.51E+06
Theoretical number of stages	*20*	40	14
Overall column efficiency	82	75	72
Actual number of stages	*26*	60	24
Annual capital cost ($/yr)	2.98E+04	3.62E+04	1.65E+04
Total annual capital cost ($/yr)		8.26E+04	
Cooling water cost ($/yr)	1.28E+04	1.17E+04	5.42E+03

steam cost ($/yr)	1.73E+05	2.04E+05	1.03E+05
Operating cost ($/yr)	1.86E+05	2.16E+05	1.09E+05
Total operating cost ($/yr)	5.10E+05		
TAC ($/yr)	2.16E+05	2.52E+05	1.25E+05
TAC of the whole sequence ($/yr)	5.93E+05		
Total annual cost saving %	6.056285485		
Total operating cost saving %	12.2753769		
Total QR (KJ/hr)	*7.02E+06*		
Total QR saving %	12.40803093		
TAC saving %	11.45905826		

Case 3: Case 3: From Table (5) below, we can notice the effects of middle component purity (0.8) for the previous scheme, Distillation Sequence with 80%midle component purity), where the reboiler duty of the second column is reduced from 3.91E+06 kJ/h to 2.45E+06 kJ/h. And 1.58E+06 to 1.48E+6 kJ/h For third column with respect with base case. These effects are translated into energy saving of 19.1 % and TAC saving of 17.74 % as shown next table

Table 5: Case 3, Middle component purity (80%)

	Column 1	Column 2	Column 3
Column pressure (kpa)	*1455*	654.9	456.9
Column diameter (m)	0.85	0.74	0.55
αave	1.960026	1.2907	2.089
Reflux ratio	4.3	5.6	1.85
Condenser duty (KJ/hr)	3.81E+06	2.93E+06	1.59E+06
Reboiler duty (KJ/hr)	2.54E+06	2.45E+06	1.48E+06
Theoretical number of stages	*20*	40	14
Overall column efficiency	82	76	72
Actual number of stages	*26*	60	24
Annual capital cost ($/yr)	3.00E+04	3.22E+04	1.68E+04
Total annual capital cost ($/yr)	7.90E+04		
Cooling water cost ($/yr)	1.28E+04	9.88E+03	5.37E+03
steam cost ($/yr)	1.74E+05	1.67E+05	1.02E+05
Operating cost ($/yr)	1.87E+05	1.77E+05	1.08E+05

Total operating cost ($/yr)	4.72E+05		
TAC ($/yr)	2.17E+05	2.09E+05	1.25E+05
TAC of the whole sequence ($/yr)	5.51E+05		
Total annual cost saving %	10.104524		
Total operating cost saving %	18.90037931		
Total QR (KJ/hr)	6.49E+06		
Total QR saving %	19.10462651		
TAC saving %	17.74583435		

Results of energy consumption TAC saving in addition to capital and operating cost saving for all the studied cases are summarized in Tables 6 and 7. The energy saving for the case 2 is reached to 12.4%while for case 3 erased to about 19.1% as shown in table 7. For there more the TAC saving for case 2 and case 3 were 12.4% and 17.74%. there are signification differences in both energy and take saving between case 2 and case 3. The saving in case 3 increased to 6.70% and 6.34% for energy and TAC saving when respectively. The results proved the importance of using middle component purity in distillation processes for high yield separation of hydrocarbons.

Table 6: TAC and total Q_R saving percentage for all the studied schemes

Studied Scheme	TAC saving (%)	Energy saving (%)
Case 1: Conventional direct distillation (base case)	0	0
Case 2: Conventional direct distillation sequence with 85% middle component purity	11.40	0 12.4
Case 3: Conventional direct distillation sequence with 80% middle component purity	17.74	19.104

The comparison of operating cost and capital cost for the studied cases are performed and the results is presented in Table 7. the operating cost of case 2 drops to 12.27%for case 3.where only drops for the capital cost of case 2 and 10.10% drops e for case 3. The decrease in the capital cost of the direct distillation processes is considered to be high ,which will affect the cost of products.

Table 7: Capital and total operating cost saving percentage for all the studied cases

Studied Scheme	Capital cost saving %	Operating cost saving %
Case 1: Direct distillation (base case)	0	0
Case 2: Direct distillation with85% middle component purity)	6.05	12.27
Case 3: Direct distillation with 80% middle component purity	10.10	18.90

Discussion

During this process, it was noticed that the largest increase in the cost of the second column is due to the largest size because of its stages which reached to 44 stages, while the other two columns reached the first 20 stages and the third 16 stages. It lead to the second column was the most energy consumption and costly in capital. This is due also to the separation of iC4 and nC4 more difficulty and knead due to the convergence of their physical and chemical properties (close to zoetrope system) and while this was not noticed in columns 1 and 3. The component C3 was separated from the mixture in the first column and nC4 was separated from iC5 in the 3th column resulted from the wide differences in their physical properties, particular boiling points which is the main factor in the distillation processes. Likewise, it was noted that the volume of coolers and boilers in the second column is larger than in the other two columns, which increased the cost of construction and operation process in terms of providing steam and pure water for the coolers.

Conclusions

The following points can be deduced from the present study

1. In distillation operations when separating a quaternary mixture, the purity of intermediate vehicles plays a significant role in cost and energy saving.
2. In the design process, we observed that the second column is always more expensive in energy and cost of capital because ISO-Butane and N-Butane . These compounds are close to physical and chemical properties.
3. Distillation process is a continuous separation process It is very important to reduce the cost of high quality products, otherwise the process will not be economically viable.
4. By reducing of the energy consumed, it means a decrease in the waste gases as a result of the combustion of light fuel or natural gas, which means environmental protection.

References

1- Perry and Chilton "Chemical engineering hand book", McGraw Hill company, 1973, 5th edition.

2- ZheyuJiang and Rakesh Agrawal, Process intensification in multi-component distillation: A review of recent advancements, Chemical Engineering Research and Design, V 147, 2019, 122-145.].

3- M. Sittig, 1978, Petrolum Refinary Industry Energy saving and Environmental Control, Noyes Data Corporation, New Jersey.]

4-Dhole, V.; Buckingham, (1994). *P. Refinery column integration for debottlenecking and energy saving, ESCAPEIV Conference, Dublin, March*.

5- Doherty, M. F. and Malone, M. F. J., Eds., (2001). Conceptual Design of Distillation Systems. McGraw-Hill Inc., New York.

6- Peter M. S., and Timmerhaus K. D., (1988). "Plant Design and Economics for Chemical Engineers".3d ed., McGraw-Hill Inc., New York.

7- Emtir M., Mizsey P., and Fonyo Z., (2003) " Economic and controllability investigation and comparison of energy-integrated distillation schemes", BudapestUniversity of Technology and Economics. Budapest, Hungary. Chem. Biochem. Eng.

8- Smith, R. (2005). Chemical Process design and integration, John Wiley and Sons Ltd, England.

9- King C .J., Separation processes 2nd Ed., McGraw-Hill Inc., (1980) New York.

10 - [Rohm and Wenzel, 2003] Wenzel, S. and Rohm, HJ (2003). Thermally coupled distillation columns using total cost optimization. Chemical Engineering & Technology.

11- Stichlmair J. G., and Fair J.R., (1998). Distillation Principles and Practices, Wiley-VCH, Canada.

12- Abolgasem Elaahi and William L. Luyben (1983). *Alternative Distillation Configurations for Energy Conservation in Four- Component Separations*, Ind.Eng.Chem. Process Des.Dev.,vol 22, No.1

13- Suphanit B., Bischert A., and Narataruksa P., Exergy loss analysis of heat transfer across the wall of the dividing-wall distillation column. Energy 2007, 32, (11), 2121- 2134.

15- Chemical Engineering magazine, (April, 2015) Chemical Engineering Plant Cost Index. Economic indicators,(www.che.com/pci)

Appendixes

Appendix A: Sizing and costing of Equipments

The sizing of distillation columns and heat transfer equipment requires the determination of flow rates, temperatures, pressures, and heat duties from the flow sheet of mass and energy balance, and these quantities can then be used to determine the capacities needed for the cost correlation .

A1; Sizing and Costing of distillation columns

1- The light and heavy key is determined, and then the relative volatility is calculated using Equation

$$\alpha_{LK/HK} = \frac{K_{LK}}{K_{HK}} \qquad [A1.1]$$

Where $\alpha_{LK/HK}$ = Relative volatility of the light key relative to the heavy key

K_{LK} = Equilibrium constant of the light key

K_{HK} = Equilibrium constant of the heavy key

Where K_{LK} , K_{HK} are taken from HYSYS simulator results.

2- The average viscosity μ_{av} between the top and bottom of the column is estimated.

3- For a given number of theoretical stages HYSYS simulator calculates column diameter after converging for selected valve tray distillation column and for flooding rate ranging from 70-80%.

4- In order to estimate the actual number of trays (N_{actual}), overall column efficiency (E_O) is estimated from Doherty & Malone correlation (2001)[5]:

$$E_o = 0.24 + 0.76\exp\left(-(\mu\ \alpha\ )^{0.5}\right) \qquad [A1.2]$$

Then the actual number of trays N_{actual} can be estimated by dividing the theoretical number of trays N in the column section by the efficiency, E_0 :

$$N_{actual} = \frac{N_{theoritical}}{E_0} \qquad [A1.3]$$

5- The tray stack height is then calculated using Equation [3-4], assuming 0.6m (24 in) tray spacing (Doherty, and Malone, 2001)[5]:

$$h = (N-1)(0.6) \qquad [A1.4]$$

Where h = tray stack height in meters

6 - By adding three meters as disengagement space (1.2m for the top, and 1.8m for the bottom of the column), the total height can be estimated:

$$H = h + 3.00 \qquad [A1.5]$$

Where $H =$ the total height of the column in meters

Capital cost of distillation columns

The capital cost of the column consists of the installation cost of the column shell and the installation cost of the tray stack.

For carbon steel construction distillation column with valve tray internals the updated bare module cost (installation cost) for each of the column shell and tray stack can be calculated by using the following cost equations that are updated with the Marshall & Swift index. For comparison a single value of M&S=1490.2 is selected, (Chemical Engineering magazine April, 2015)[14].

Installed cost of column shell,

$$\$ = \left(\frac{M\&S}{280}\right) * (2982) * d^{1.066} * H^{0.802} \qquad [A1.6]$$

If the design pressure (P) is more than 345 kpa, a correction factor [1+1.45*10^(-4)(P345)] is applied.

Installed cost of the column trays,

$$\$ = \left(\frac{M\&S}{280}\right) * (136.14) * d^{1.55} * h \qquad [A1.7]$$

Where ; d is the column diameter, H is the column height and h is tray stack height

The total column cost is the sum of the installed cost of column shell and the installed cost of column trays.

Column cost = (Installed cost of shell) + (Installed cost of trays) [A8]

A2; Sizing and Costing heat transfer equipments

The heat transfer area , of the condensers, reboilers and heat exchangers are calculated assuming the overall heat transfer coefficients, U, given in Table A1. and are calculated according to the following equation:

$$A = \frac{Q}{U \cdot LMTD} \qquad [A2.1]$$

where Q is the heat duty, and $LMTD$ is the logarithmic mean temperature difference:

$$LMTD = \frac{\Delta T_2 - \Delta T_1}{\ln \dfrac{\Delta T_2}{\Delta T_1}} \qquad [A2.2]$$

ΔT_2 is the temperature difference between the inlet streams, and ΔT_1 is the temperature difference between the outlet streams of the heat transfer equipment.

Table A1: Overall heat transfer coefficients: (Peters &Timmerhaus 1988)[6]

Heat transfer equipment	Overall heat transfer coefficient, *U*, [kJ / m^2 h °C]
Condenser	2800
Reboiler	3400
Heat exchanger	2100

Capital cost of heat transfer equipment

The cost of heat transfer equipment can be correlated as a function of surface area. Assuming shell and tube, floating head, and carbon steel construction, can be calculated by using the following cost equations that are updated with the Marshall & Swift index.

Installed cost of the heat equipment, (ICH) ,

$$\$=(\frac{M\,\&\,S}{280})*(1562)*A^{0.65} \qquad [A2.3]$$

where: A = area m2, 18.6 < A < 464.5 and design pressure up to 1034.2 kpa.

Appendix B ;Estimation of operating costs

The operating cost includes the utility costs which is calculated per year as a function of the operating hours.

The operating costs are assumed only steam and cooling water costs. So we take the operating hours per year is set to be 8000 (hr).

The flow rate of cooling water, $\dot{m}_{CW}$, is calculated using the following equation (Turton et.al, 1998):[17]

$$\dot{m}_{CW}=\frac{Q}{C_P.\Delta T} \qquad [B.1]$$

Where the specific heat capacity of water, C_P, equals 4.181 J/K·g and ΔT is temperature difference.

Cooling water cost = $\dot{m}_{CW}$ × cooling water price × operating hours ×10^{-3} [B2]

The flow rate of steam, $\dot{m}_{STM}$, is calculated using the following Equation:

$$\dot{m}_{STM}=\frac{Q}{\lambda} \qquad [B3]$$

Where λ is the latent heat of steam, equal to 2070 kJ/kg for low-pressure steam and 2000 kJ/kg for medium-pressure steam, (Emtir, et.al 2004).[7]

Steam cost = $\dot{m}_{STM}$ × LP-steam price × operating hours ×10^{-3} [B4]

The utility prices used to calculate operating costs are given in Table 3.5:

Table B1: Utilities cost data(Emtir, et.al 2004)[7]

	utility prices Based on European prices	
Utility	Temperature (°C)	Price ($/ton)
LP-steam	160	17.7
MP-steam	184	21.8
Cooling water	23-40	0.0272

Appendix C: Estimation of Total Annual Cost (TAC)

The capital and operating costs are annualized over a period referred to as the plant lifetime, and assumed to be 10 years (8000 hrs / year) and Operating costs were assumed just only the utility cost (steam and cooling water).

Annual Capital Cost =Capital cost /Plant life [C1]

the TAC (includes capital and utility costs) can be calculated according to the following equation:

TAC = Annual operating cost + Annual capital cost [C2]

The economic study provides the TAC as indicator that characterizes the different distillation design alternatives.