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CHP GENERATION IN HEAT INTEGRATED SEPARATION SYSTEM

Hasanzadeh. K.* Salehi G.R. and Amidpour M.
*Author for correspondence
K.N.Toosi University of Technology, Integration Lab.
Tehran, Vanak Square
Iran, Islamic Republic of
E-mail: k.hasanzadeh@yahoo.com

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. In this paper, the possibility of producing power in a sequence of heat integrated columns in a chemical process is inspected. The aim is to find the best sequence with heat integration and its feasible power production as a means of achieving a minimum TAC (Total Annual Cost). First, all sequences have been simulated with HYSYS and then, heat integrations and feasible power production in each sequence are examined. After finding the best values for parameters of each sequence, all sequences are compared and finally the optimum state for the process is found.

INTRODUCTION

Separation of components of multi-component flows using simple distillation columns can be carried out 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 causes a considerable reduction in heat consumption of these column sequences [1]. There are a number of different methods for heat integration of columns and numerous works have been carried out about this subject so far [1-9]. Another possible optimization for columns is increasing 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 temperatures columns can be directly consumed by cryogenic systems. Numerous works of research have been done so far related to the exergetic examination of columns and sequences [10-12]. 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% [13]. Using the capability of power production between the columns greatly increases the exergetic efficiency of the process.

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

We use Kotas's relationship [14] to calculate the exergetic efficiency (ψ) .

$$\psi = \frac{\sum \Delta E_{out}}{\sum \Delta E_{in}} \tag{1}$$

This relationship is based on the ratio of the required output of the process to the input necessary to obtain that output. The exergies of flows have also been obtained using the following relationship.

$$E = E_{ph} + E_{ch} \tag{2}$$

Ambient temperature and pressure are respectively 30oC and 1atm.

In order to separate multi component flows to n product groups, the number of column sequences (Sn) is calculated using King's relationship [15]:

$$S_n = \frac{[2(n-1)]!}{n!(n-1)!} \tag{3}$$

CASE STUDY

In this paper, flow separation of compounds such as propane, propylene, butane, butylene, pentane, and hexane to four groups of products is examined. The possible sequences for this process are as follows: Distributed sequence (figure 1): AB/CD; A/B; C/D. Direct sequence (figure 4): A/BCD; B/CD; C/D. Indirect sequence (figure 9): ABC/D; AB/C; A/B. Direct-

indirect sequence (figure 7): A/BCD; BC/D; B/C. Indirect-direct sequence (figure 11): ABC/D; BC/D; B/C.
The aim is to find the best arrangement with its heat integrations and power production in order to achieve the minimum value of TAC.

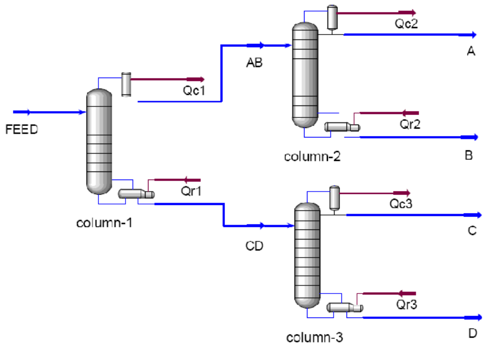

Fig. 1. Primary Distributed sequence

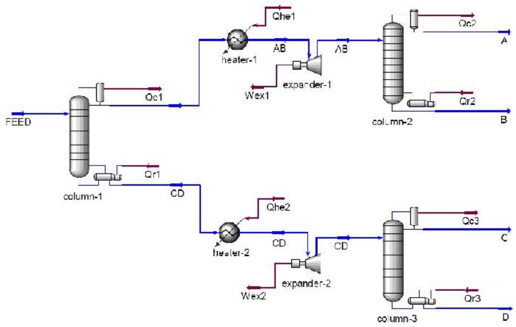

Fig. 2. Distributed (1X2&1X3) sequence

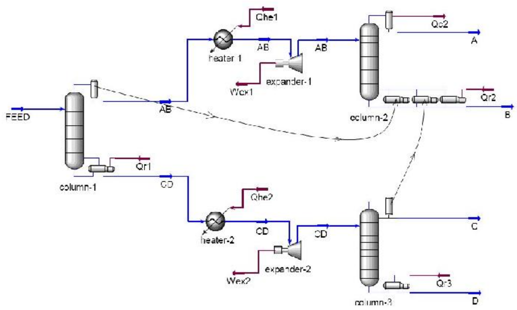

Fig. 3. Distributed (C1&C3-R2)(1X2&1X3) Sequence

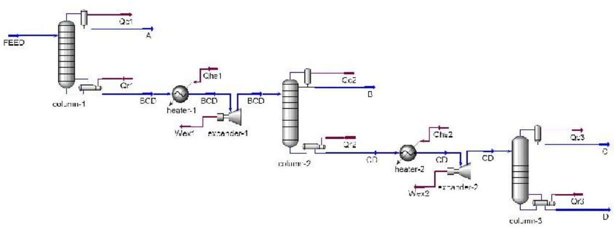

Fig. 4. Direct (1X2&2X3) sequence

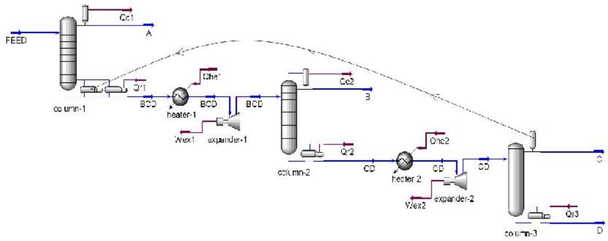

Fig. 5. Direct (C3-R1)(1X2&2X3) sequence

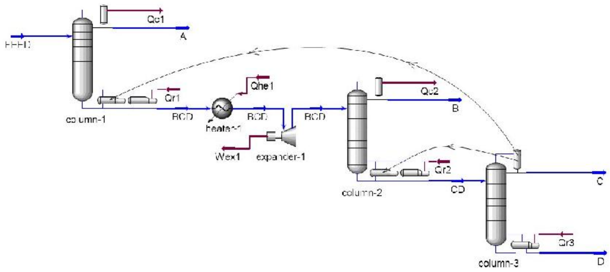

Fig. 6. Direct (C3-R2&R1) (1X2) sequence

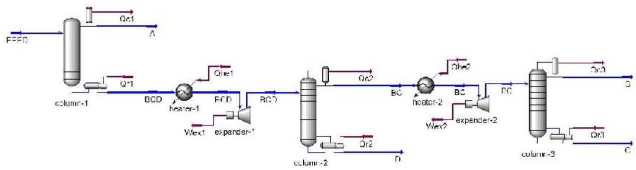

Fig. 7. Direct-Indirect (1X2&2X3) sequence

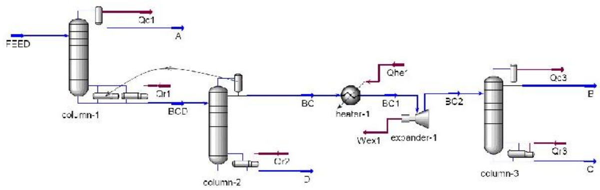

Fig. 8. Direct-Indirect (C2-R1)(2X3) sequence

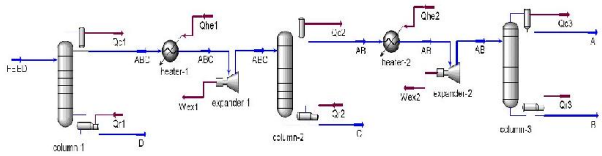

Fig. 9. Indirect (1X2&2X3) sequence

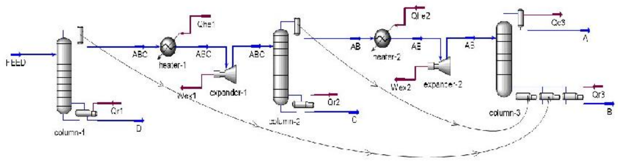

Fig. 10. Indirect (C1&C2-R3)(1X2&2X3) sequence

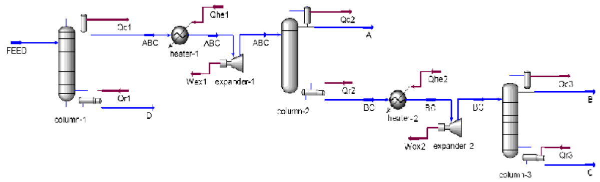

Fig. 11. Indirect-Direct (1X2&2X3) sequence

DISCUSSION ON THE RESULTS

The distributed sequence of the initial process is chosen as the base for comparison of other distributions (figure 1). The exergetic efficiency of the initial process is 0.28

In the sequence Distributed(1X2&1X3), power is produced using the pressure difference between columns 1 and 3 and also temperature increase and enthalpy increase of the CD flow using HP steam and passing it through the expander(figure 2). Operational costs in this state decreased by 44.03% relative to the base case. The heat consumptions of columns 2 and 3 have decreased greatly compared to the case without the expander as a result of the existence of the steam flow, while the heat outputs of condensers have increased. The total steam cost has increased because of using HP steam and high heat consumption in heat exchangers heating the flow entering the expanders. The scrutiny resulting from minimization of the TAC function showed that using HP steam for heating the intake flow of the expanders in different sequences is preferred to LP and MP steams. Reflux ratio and diameters of columns 2 and 3 have increased compared to the case without the expander which is a result of increasing the steam volume of columns. Although the costs related to steam and cold water have increased in this case, but the operational costs have shown a high improvement which is a result of the benefits involving power production of expanders. In the distributed state 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. Capital costs have in turn increased by 15.57% because of high costs of expanders. Finally, TAC has decreased by 37.7% compared to the base case. The exergetic efficiency is 97% 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. Since the intake flow rate to column 2 has a high value, the expander put in its way has produced a much higher power compared to the second expander.

In the case Distributed (c1&c3-r2) (1X2&1&3) power is produced by putting two expanders in the way of AB and CD flows (figure 3). Saturated liquid streams of column 1 have been superheated 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.

The operational cost have improved by 69.32% compared to the base case which is a result of the optimization caused by two heat integrations and two expanders installed in the process.

Description	Primary Distributed	Distributed (1X2&1X3)	Distributed (C1&C3-R2) (1X2&1X3)
Total actual plates	260	273	272
TAC ($/yr)	7177384.24	4471592.92	2833536.17
Op. cost saving (%)	-	44.03	69.32
Capital saving (%)	-	-15.57	-13.53
TAC saving (%)	-	37.7	60.52
TAC sav. by heat (%)	-	-	27.86
TAC sav. by power (%)	-	37.7	32.66
Thermo. Eff.	0.28	0.97	0.89

Description	Direct (1X2&2X3)	Direct (C3-R1) (1X2&2X3)	Direct (C3-R2&R1) (1X2)
Total actual plates	264	263	270
TAC ($/yr)	5012322.9	4804199.55	5452067.48
Op. cost saving (%)	34.31	37.77	27
Capital saving (%)	-4.69	-6.51	-0.92
TAC saving (%)	30.17	33.06	24.04
TAC sav. by heat (%)	9.85	13.14	11.1
TAC sav. by power (%)	20.32	19.92	12.94
Thermo. Eff.	0.62	0.59	0.4

Description	Direct-Indirect (1X2&2X3)	Direct-Indirect (C2-1)(2X3)	Indirect (1X2&2X3)
Total actual plates	264	265	294
TAC ($/yr)	4772868.12	5623837.03	4362824.66
Op. 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 sav. by heat (%)	1.2	6.48	-17.31
TAC sav. by power (%)	32.3	15.17	56.52
Thermo. Eff.	0.73	0.43	0.9

Description	Indirect (C1&C2-R3) (1X2&2X3)	Indirect-Direct (1X2&2X3)
Total actual plates	303	292
TAC ($/yr)	2190632.52	4615472.59
Op. cost saving (%)	84.39	42.76
Capital saving (%)	-55.96	-23.77
TAC saving (%)	69.48	35.69
TAC sav. by heat (%)	17.15	-8.11
TAC sav. by power (%)	52.33	43.8
Thermo. Eff.	0.85	0.8

Table 1: Optimal schemes

The exergetic efficiency in this state is 0.89. Condenser heats of columns 2 and 3 have increased and their reboiler heats has 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 compared to the state without expander due to increase of steam volume. The initial cost has shown a 13.53% increase 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 improved by 60.52% which involves a thermal improvement of 27.86% and the optimization caused by power production of 32.66%. In this case also the expanders have produced a high power due to high pressure difference and high flow rate between the columns and super heating of the intake flow of expanders.

In the case Direct (1X2&2X3) a large power has been produced using the pressure difference between the columns and employing expanders between them (figure 4). In this case the operational cost has decreased by 34.31%. The initial cost has increased by 4.69% and TAC has improved by 30.17 %
TAC has had a heat optimization of 9.85% and the optimization involving power production has been 20.32%. The exergetic efficiency of this case is 0.62. In this case also the intake flows of expanders have been heated using the HP steam. In this sequence, since light components exit from the tops of columns, column pressure reduces sequentially and therefore creating a reasonable pressure difference between the columns for the purpose of producing power has become possible. However, since products A and B have left the process in pure states in primary columns, a high potential of high pressure for power production has been lost, because products A and B constitute 45% and 29% of the intake flow rate of the process. The fact that the columns are sequential in this case has resulted in the pressure difference between two columns influencing the pressure difference of the two other columns. For instance, by increasing 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, assuming that column 1 has the highest pressure and column 3 has the lowest pressure. The reflux ratio and diameters of columns 2 and 3 has also increased compared to the case without the expander as a result of increasing of the volume of steam in expanders. Flow rates between columns are usually higher in sequences leaving the expanders consecutively compared to the distributed case because in the distributed case the intake flow rate of the process is divided into different parts and flows between the columns while in the consecutive state the intake flow rate gradually reduces throughout the process and therefore the total flow rate between the columns us usually more than the intake of the process. The negative point about the consecutive state is the high heat consumption of columns because the air intake flow rate of columns is high and we have limited freedom to carry out pressure changes in columns for the purpose of heat integration.

In the Direct (C3-R1) (1X2&2X3) case, a considerable amount of power is produced using the pressure difference between the columns and employing two expanders in the process. Heat integration is also performed between the condenser of column 3 and the reboiler of column 1(figure 5). The operational costs in this case have decreased by 37.77% The intake flows to the expanders have been superheated by the HP steam. The intake steam to columns2 and 3 has decreased the heat of reboilers and increased the heat of condensers. In spite of heat integration and power production in this state, the improvement in operational costs is almost equal to the case Direct(1X2&2X3) which is a result of reduction of power production as a result of pressure changes carried out with the purpose of heat integration between the columns, as in this state the pressure of column 3 has been increased to increase the temperature of its condenser and transfer heat to the reboiler of column 1 and the pressure of column 1 has also been decreased to decrease the temperature of its reboiler. The aforementioned factors have considerably reduced the power production potential between the columns. On the other hand, as a result of high temperature difference between the columns, we have less freedom to change their pressures because it rapidly increases the temperatures of condensers and reboilers beyond the allowable limits. The reason for this fact is the existence of materials with different volatilities in the columns. The exergetic efficiency in this case is 0.59 which is lower than the case Direct (1X2&2X3) . The capital costs increased by 6.51% and TAC improved by 33.06%, its heat optimization contribution and power production contribution being 13.14% and 19.92%, respectively. Increasing the volume of steam in columns 2 and 3 increases the reflux ratio and the diameter of these columns.

In the case Direct (C3-R2&R1) (1X2), since there is heat integration, the pressures of columns allows only one expander in the process, which is located between columns 1 and 2 (figure 6). The output flow form the lower part of column 1, is superheated using the HP steam. Here we have two heat integrations between the condenser of column 3 and reboilers of columns 1 and 2. From results, it is evident that the heat transferred through heat integration in this case is lower than that of the case Direct (C3-R1) (1X2&2X3) . Operational costs have decreased by 27% which represents a lower improvement compared to Direct(1X2&2X3) and Direct(C3-R1)(1X2&2X3) cases since the power production in this case has had a high decrease greatly reducing the exergetic efficiency of this case to a very modest value of 0.4. Capital costs have increased by 0.92% compared to the base case and TAC has improved by 24.04%, its heat optimization and power production contributions being 11.1% and 12.94%, respectively. The reflux ratio and the diameter of column 2 have also increased as a result of increased volume of the steam in the column.

In the case Direct-Indirect (1X2&2X3), two expanders have been put between the columns (figure 7). Operational costs have reduced 39.32% compared to the base case. The expanders' intake flows have been superheated using the HP steam to increase the power production of expanders. The reason of power production in this case compared to the case Direct (1X2&2X3) is the separation of product B in column 3 since product B constitutes 29% of the flow rate of the total intake flow and this much flow leaving column 2 in the case Direct (1X2&2X3), have largely reduced the power production between columns 2 and 3. The exergetic efficiency of this case is 0.73 which represents high power production in the process. Capital costs have increased by 15.47% and TAC have decreased by 33.5%, heat improvement and power production contributing to 1.2% and 32.3%, respectively.

In the case Direct-Indirect (C2-R1) (2X3) it is only possible to put one expander between the columns 2 and 3. Heat integration has also been carried out between the condenser of column 2 and the reboiler of column 1 (figure 8). In this case operational costs have reduced by 25.8% relative to the base case. In spite of its heat integration, this case has had a lower improvement in operational costs compared to the case Direct-Indirect (1X2&2X3) which is a result of lack of an expander between columns 1 and 2, due to the positive pressure difference between them, greatly reducing power production. The exergetic efficiency of this case is 0.43. Capital costs have increased by 13.34% and TAC has improved by 21.65%, thermal optimization and power production contributing 6.48% and 15.17%, respectively. In this case also steam intake flow and higher steam volume in columns 2 and 3, has increased their diameters and reflux ratios.

In the case Indirect (1X2&2X3) two expanders have been put between the columns (figure 9). Operational costs have decreased by 48.47%. Capital costs increased by 41.24% and TAC reduced by 39.21% , 17.31% of the decrease in TAC was caused by changing the sequence and 56.52% of it was caused by the power produced by expanders. The exergetic efficiency of this case is 0.9 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 compared to the costs imposed by more heat consumption.

In the case Indirect (C1&C2-R3) (1X2&2X3), like the former case, two expanders have been put between the columns and the intake flow of expanders have been superheated using the HP steam (figure 10). 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 reduced by 84.39%. Capital costs have been augmented by 55.96% and TAC has lowered by 69.84%. The improvement cause 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 are high flow rates going through the expanders put between the columns. The exergetic efficiency of this case is 0.85.

In the case Indirect-Direct, a high power is produced by putting two expanders between the columns (figure 11). The operational costs have decrease by42.76% compared to the base case, capital costs have increased by 23.77% and TAC has improved by 35.69%. For the thermal part, TAC has increased by 8.11% and for the power production part, it has reduced by 43.8%. Since product A has left the column in column 2, the power production of the second expander has had a considerable reduction in this case compared to the case Indirect (1X2&2X3). The exergetic efficiency of this case is 0.8. The reason for high increase of capital costs in indirect cases as compared to the other cases is the high costs of expanders.

CONCLUSIONS

In this study, all possible states of heat integration and installation of expanders between the columns in order to produce power in different sequences for separation of a four component flow are studied, considering the constraints of the process. 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 such that in this sequence, not only 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. Separation of products A and B in column 3 has increased the possibility of heat integration in this 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 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 which increases the costs related to heat consumption of columns compared to Distributed cases. Placing expanders between the columns, not only produces power, but also it brings about intake steam 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.

A high temperature difference inside the columns is one of the factors making changing their pressures difficult. The reason for that is having 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. 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 high freedom in creating heat integration, but also it should include high flow rate flows between the columns. 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

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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)

$$H = (N-1)\times 0.6 + 6.0 \qquad \text{(A.2)}$$

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* < 464.5m2.

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