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A design procedure for heat-integrated distillation column sequencing of natural gas liquid fractionation processes

Hanareum Yoo, Michael Binns, Mun-Gi Jang, Habin Cho, and Jin-Kuk Kim†

Department of Chemical Engineering, Hanyang University, 222, Wangsimni-ro, Seongdong-gu, Seoul 133-791, Korea
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Abstract–The separation of NGL (natural gas liquids) in gas processing is energy-intensive, requiring systematic process design and optimization to reduce energy consumption and to identify cost-effective solutions for the recovery valuable hydrocarbons. As NGL fractionation processes require a sequence of distillation columns to separate multi-component mixtures the determination of optimal energy-efficient distillation sequences and operating conditions is not a simple task. A design methodology is proposed in this study in which the process simulator Aspen HYSYS® is linked with an optimization algorithm available in MATLAB®. The proposed methodology involves a procedure where in the first step possible distillation sequences are screened using a short-cut distillation column model. In the second step a few selected and promising candidate distillation sequences are further simulated and optimized, again using the same short-cut model. Finally, rigorous simulations are used to validate and confirm the feasibility of the optimal designs. A case study is presented to demonstrate the applicability of the proposed design framework for the design and optimization of NGL fractionation processes in practice.

Keywords: Distillation Sequencing, Optimization, Energy Recovery, Natural Gas Separation

INTRODUCTION

The composition of natural gas includes methane, heavier hydrocarbons (ethane, propane, butane, etc.) and other impurities. Valuable natural gas liquid (NGL) products can be extracted from natural gas through various separation processes to give products such as LPG (Liquefied petroleum gas) while meeting the product specifications of sale gases. The NGL fractionation process is the most widely-used method for the separation of NGL in gas processing in which a number of distillation columns are used to generate products including ethane, propane, butane and C5+. However, distillation processes are energy-intensive [1,2] and hence improving the energy efficiency of NGL separation is one of the main targets for the reduction of costs [3].

NGL fractionation processes are used to separate a multi-component mixture through the application of a series of columns in sequence. Hence, the design of NGL fractionation processes is very complex because of the large numbers of possible distillation sequences, design interactions and key design variables. This complexity is significantly increased for NGL fractionation when heat integration is simultaneously considered for the purpose of improving system-wide energy efficiency.

Heat integration is one of the most widely-used methods for improving energy efficiency which has been successfully applied to a wide range of process industries over the last three decades. Heat integration methodology aims to maximize heat recovery in the processes by considering thermodynamic principles [4,5].

†To whom correspondence should be addressed.
E-mail: jinkukkim@hanyang.ac.kr

Also, in addition to the basic column design (including a single reboiler and condenser) a number of studies have looked at various complex column designs which can potentially reduce energy consumption or simply reduce the number of columns required (possibly reducing the capital costs and the space required for equipment). For example, Errico and Rong [6] show that two columns with a connecting liquid or vapor side stream can be used to separate a four component mixture rather than a conventional sequence with three simple columns. Additionally, Long and Lee considered various column modifications including the use of integrated heat pumps [7], side reboilers and side condensers [8], columns with dividing walls [9,10] and the use of thermally coupled columns [8,11]. However, while such methods have a number of benefits, they also increase the complexity of the column design procedure (simulation and optimization) in addition to increasing the complexity of any implementation. Hence, this study focuses on the development of designs using sequences of basic columns.

When evaluating the performance of different distillation column sequences and operating conditions, rigorous models are considered to be computationally demanding for the purpose of optimization. Therefore, the application of short-cut models is a practical way to evaluate various sequences in a quick and efficient manner. The FUG (Fenske-Underwood-Gilliland equation) method is a well-known short-cut model for the design of distillation columns [12]. The FUG method can be used as a basic tool to rapidly assess design options and their techno-economic impacts and to provide an initialization for the simulation of rigorous distillation columns [4,13].

Various design methodologies have been proposed to identify the optimal distillation sequences for multicomponent mixtures. For example, Shah and Kokossis [14] and Caballero and Grossmann [15] proposed optimization approaches for identifying the optimal configuration by considering multiple different configurations simultaneously inside superstructures. Additionally, optimization of distillation sequencing simultaneously considering heat integration has been carried out by An and Yuan [16] and Jain et al. [17].

Although methods such as these generally use short-cut models, the resulting formulation and the construction of the superstructure model is still relatively complex. Additionally, these methods typically aim to generate a single "best" configuration and set of operating conditions and they do not consider any alternatives. In particular, an NGL plant may be required to operate at a range of operating conditions around the optimal due to practical limitations (e.g. engineering limits), and so it is beneficial to consider the performance at different conditions.

Therefore, we propose a design methodology that involves a screening step where different possible configurations and operating conditions are evaluated followed by optimization of a small number of promising options. This approach avoids the complexities associated with the construction and simulation of superstructures. Additionally, this method generates results at a range of different conditions, which provides a more complete picture showing how the systems perform at optimal and sub-optimal conditions. The screening and optimization are performed using a short-cut model, and the final solutions are validated with more rigorous simulations to confirm their performance characteristics at the optimal points.

This paper first examines design issues including column sequencing and heat integration. The proposed design methodology is then presented, which explains how to screen the distillation sequences in a holistic manner, how to incorporate heat integration and how these options can be optimized to find the best values of key design variables. Finally, an NGL fractionation case study is presented to illustrate how heat-integrated distillation sequences can be designed and optimized in practice and to demonstrate the applicability and usefulness of the design method proposed in this study.

PROCESS DESIGN

1. Design Issues Associated with Distillation Sequencing for NGL Fractionation

In this study simple distillation columns containing one feed and two product streams (in addition to one reboiler and one condenser) are used for NGL fractionation. It is not a simple task to determine the optimal sequence of distillation columns for the separation of multicomponent mixtures due to the large number of possible and feasible distillation sequences (the number of possible sequences increases exponentially with the number of products). Fig. 1 shows five possible sequences for the separation of a four component mixture.

For distillation sequencing one common engineering practice is to apply heuristic rules which can be used by engineers to guide decision making and to reduce the number of distillation sequences to be studied in detail [4,18]. Although heuristic rules can be applied without difficulty, there are often cases in which the heuristic rules conflict with each other. These heuristics are only empirical and hence they are not always applicable for multi-component distillation systems. For these reasons, a more systematic method should be used to screen different sequences and to find the optimal sequence subject to product specifications, process economics and design constraints. However, while the use of heuristics can be considered an over-simplification, the use of systematic superstructure-based optimization [14,15] may be considered to be unnecessarily complicated, requiring the simultaneous simulation and optimization of multiple different options in parallel. Hence, considering these issues a sequential strategy is employed here which uses systematic screening of different configurations and operating conditions followed by optimization of the most promising options. In this way the most appropriate configurations can be identified without relying on heuristic rules and without the complexities associated with the development of superstructures.

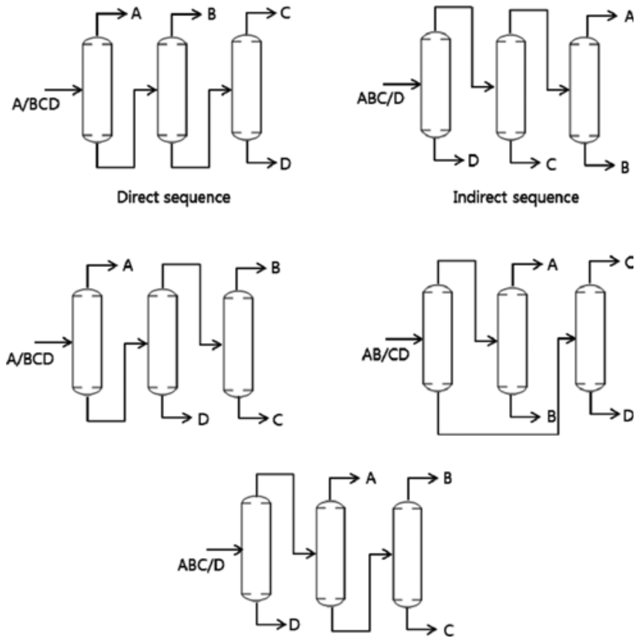

Fig. 1. Possible distillation sequences for a four component mixture.

To reduce the consumption of utilities in the columns heat integration can be applied to identify energy efficient modifications. This is possible through heat recovery in the distillation systems, connecting cold streams in reboilers with hot streams in condensers within the various columns. To analyze system-wide heat recovery energy composite streams (ECSs) and grand composite curves (GCCs) can be used, subject to the ΔT_{min} (minimum temperature difference for heat exchange, °C). GCCs can be utilized together with the temperatures of utilities provided to target and minimize duties. These graphical methods are very effective for the purpose of achieving energy savings, leading to a reduction in the total annualized cost (TAC) and providing strategies which can identify the most economic utilization of utilities.

Fig. 2 illustrates how heat integration within two columns can be implemented. The condenser of the first column provides the heat for the reboiler of the second column in Fig. 2(a), while the condenser of the second column provides the heat for the reboiler of the first column shown in Fig. 2(b).

2. Design Framework for Distillation Sequencing of NGL Fractionation

The FUG (Fenske-Underwood-Gilliland) method is used for

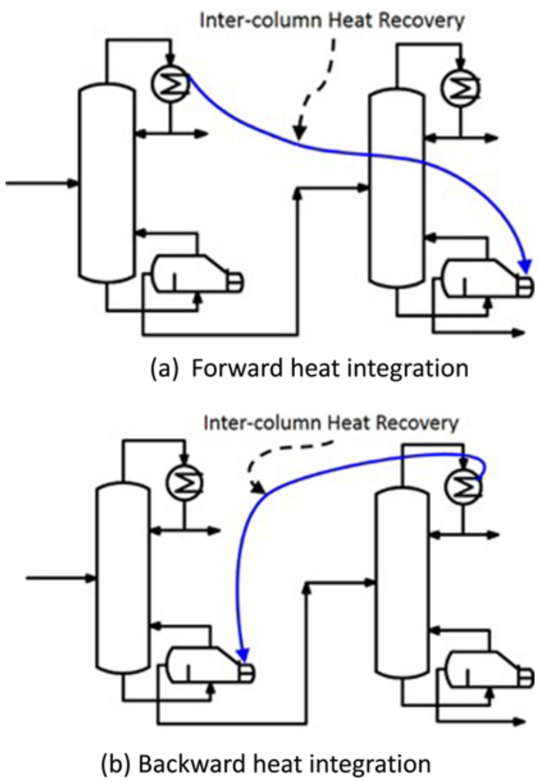

Fig. 2. Heat integration opportunities for a sequence of two simple distillation columns [4].

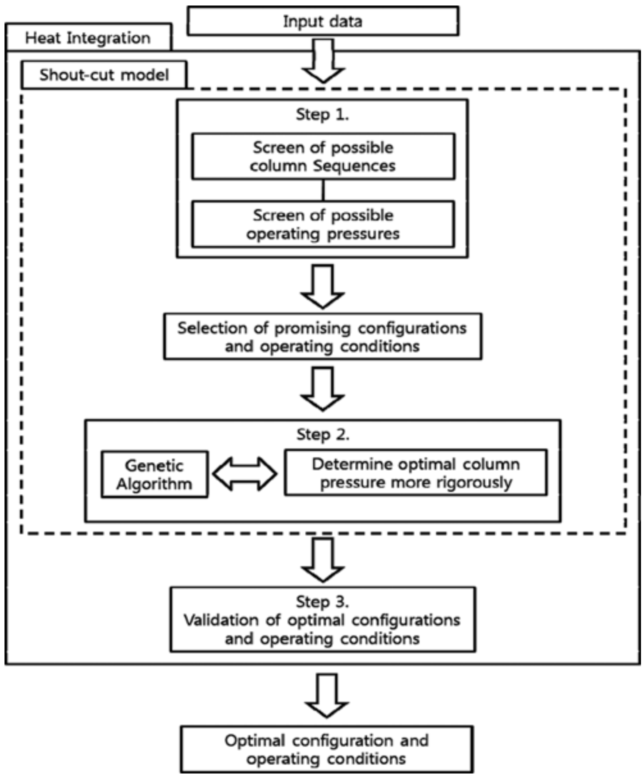

Fig. 3. Overall screening and optimization framework for distillation sequences.

the preliminary design of distillation columns; details of this short-cut method can be found in the literature [19]. For economic costing of distillation sequences, the column diameters are calculated to estimate the capital costs of distillation columns, while energy duties for reboilers and condensers are obtained to estimate energy costs (the main element in the operating costs of columns).

A key design variable for distillation columns is the operating pressure of a column, as variation in column pressure significantly affects a number of important design parameters including relative volatilities, vapor loads and temperature profiles inside the columns [20,21]. For NGL fractionation it is necessary to simultaneously select both the column pressures and the distillation sequence. To identify optimal operating conditions and column sequences for NGL fractionation a systematic and integrated design approach is required. This study proposes a design methodology based on three steps with the strategic use of a short-cut model together with optimization methods.

The procedure for screening and optimization of column sequences is shown in Fig. 3. In the first step all the possible distillation sequences are generated and the short-cut model is used to simulate the various distillation columns in each configuration. As economic performance of distillation sequences strongly depends on the choice of column operating pressure a few levels of column operating pressure for each column are selected and these different possibilities are systematically evaluated. This implementation is relatively straightforward because of the use of a short-cut model which simplifies the method. When the TAC for each sequence is calculated the maximized heat recovery potential is exploited within the sequence using energy composite streams. Based on the calculated TAC values a number of promising sequences are chosen for further evaluation. In this study the best two sequences are selected from the first step (although in principle any number of potential options can be considered at the designer's discretion).

In the second step a GA (genetic algorithm) optimization method in MATLAB® is employed to determine optimal column pressures more rigorously. As only a few levels are pre-specified for potential column pressures in the first step, optimization can identify improved solutions for the sequences selected in the first step. This second step also uses the short-cut model for simulation of distillation columns and employs heat integration to maximize heat recovery. The first and second steps are programmed within the MATLAB® environment, while thermodynamic information and physical properties required for the short-cut model computer program are provided from Aspen HYSYS®. The objective function for the optimization is to minimize the total annual cost as shown in Eq. (1).

$$\text{Minimize } C_{TAC}=C_{OP}+C_{FA}\cdot C_{CAP} \tag{1}$$

where C_{TAC}, C_{OP}, C_{CAP} and C_{FA} are the total annualized costs (USD·yr^{-1}), operating costs (USD·yr^{-1}), capital costs of columns (USD) and annualization factor (yr^{-1}). In the third and final step the selected sequences from the second step are rigorously simulated using the process simulator Aspen HYSYS® which confirms the feasibility of optimal column designs.

The GA optimization method used here is effective for the purpose of dealing with highly non-linear problems and is able to obtain optimal solutions starting from any reasonable initial values through

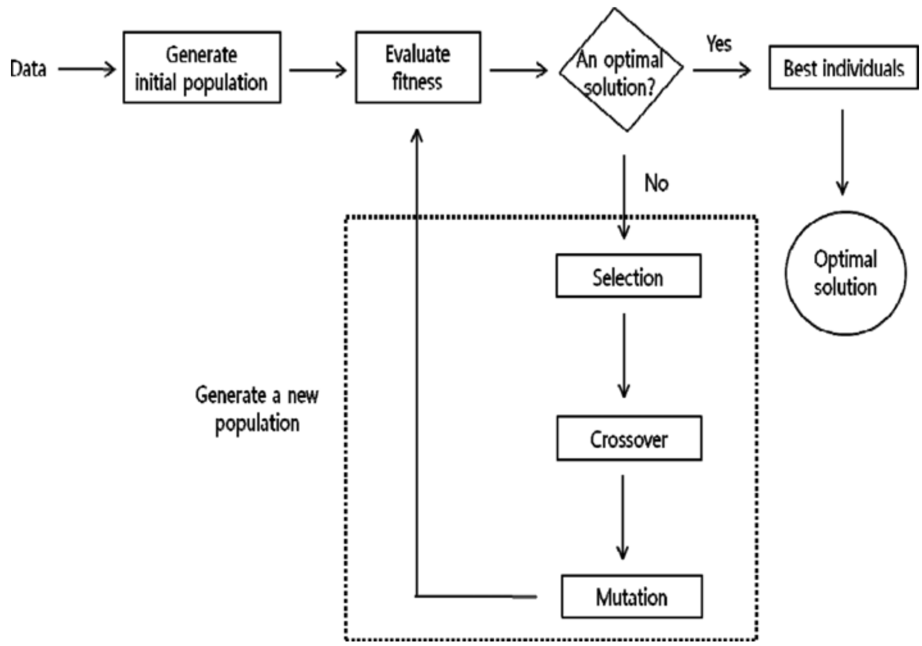

Fig. 4. Genetic algorithms (GA) procedure [22].

repeated stochastic optimization moves. Fig. 4 shows a standard GA procedure for the identification of optimal solutions.

CASE STUDIES

For the case studies considered here, ranges of operating pressures for case A (5-33 bar) and for case B (4-19 bar) are chosen so that very high and very low temperature conditions can be avoided. To evaluate economic performance for each sequence considered, the total annualized cost is calculated and details of economic costing are given in the Appendix. Also, as mentioned above, a shortcut model based on the FUG method is applied in the first and second steps of the method, while rigorous simulation using Aspen HYSYS® is carried out in the third step using the Peng-Robinson equation of state.

1. Base Case

Two cases are considered here: Case **A** involves the separation of ethane (C2), propane (C3), iso-butane (iC4), n-butane (nC4) and iso-pentane (iC5+) components with four distillation columns. Case **B** involves the separation of propane (C3), isobutene (iC4), n-butane (nC4), iso-pentane (iC5) and n-pentane (nC5+) components. Feed conditions and product specifications for cases A [23] and B [4] are given in Tables 1 and 2. The different utilities available and their costs are given in Table 3.

The following assumptions are made here:

• Feed and product streams of the distillation columns are saturated liquids.

• The reflux ratio in distillation columns is constant at a value of $1.1xR_{min}$

• Total condensers and reboilers are used.

Table 1. Feed conditions and product specifications of case A [23]

	Feed (kmol/h)	Ethane	Propane	iso-Butane	n-Butane	Pentanes
Methane	61.9	1.36				
Ethane	2,901.1	95.14	7.32			
Propane	1,980.3	3.50	90.18	2.00		
i-Butane	461.4			96.00	4.50	
n-Butane	984.4			2.00	95.00	
Butanes			2.50			3.00
i-Pentane	286.4					33.13
n-Pentanes	202.5					23.52
Pentane					0.50	
n-Hexane	203.9					26.90
n-Heptane	90.9					13.45
Temp. (°C)	29.4					
Pressure (kPa)	4,238					

Table 2. Feed conditions of case B [4]

Component	Flowrate (kmol·h^{-1})
Propane	30.3
i-Butane	90.7
n-Butane	151.2
i-Pentane	120.9
n-Pentane	211.7
n-Hexane	119.3
n-Heptane	156.3
n-Octane	119.6
Pressure (kPa)	1,400
Feed conditions	Saturated liquid

Table 3. Utility data [25,26]

Utility	Temp. (^{o}C)	Cost (USD·MWh^{-1})
Refrigerant 1 (propane)	−20	34.75
Refrigerant 2 (ethane)	−70	82.48
Cooling water	35	5
Low pressure steam (LP)	133.5	10.78
Middle pressure steam (MP)	198.6	13.25
Electricity		50

- Constant heat transfer coefficients for using steam, cooling water, propane, ethylene and process-to-process are 1, 0.8, 0.65, 0.65 and 0.65 kW·m^{-2}·^{o}C^{-1} respectively.
- Pressure drop is neglected.
- Minimum approach temperature for heat recovery is 10 ^{o}C.

The short-cut model used in this study is validated here through comparison of simulation results against design data from the literature [23], including product purity, hot utility consumption and column size as shown in Table 4. The short-cut model used here produces results similar to those in the literature, and so it is reasonable to use this model for the optimization of distillation column sequencing and operating conditions.

Table 5. Results of costing based on short-cut modeling of case A

	Sequence	TAC (USD·yr^{-1})
1	A/BCDE B/CDE C/DE D/E	2.8804×10^{7}
2	A/BCDE B/CDE CD/E C/D	2.7570×10^{7}
3	A/BCDE BC/DE B/C D/E	3.0484×10^{7}
4	A/BCDE BCD/E B/CD C/D	2.9459×10^{7}
5	A/BCDE BCD/E BC/D B/C	3.2920×10^{7}
6	AB/CDE C/DE A/B D/E	3.6845×10^{7}
7	AB/CDE CD/E A/B C/D	3.5918×10^{7}
8	ABC/DE A/BC B/C D/E	4.0131×10^{7}
9	ABC/DE AB/C A/B D/E	4.7466×10^{7}
10	ABCD/E A/BCD B/CD C/D	3.8374×10^{7}
11	ABCD/E A/BCD BC/D B/C	4.1632×10^{7}
12	ABCD/E AB/CD A/B C/D	4.6821×10^{7}
13	ABCD/E ABC/D A/BC B/C	5.2167×10^{7}
14	ABCD/E ABC/D AB/C A/B	5.9928×10^{7}

2. Case A

Application of the first step in the design procedure generates and screens fourteen different sequences for case A and two of these sequences (Sequences 1 and 2) are selected as promising arrangements based on the TAC values given in Table 5. Following the second step of the procedure these two selected sequences are optimized using GA, giving the results shown in Tables 6 and 7. Fig. 5 shows the heat recovery scheme in sequence 1 where the condenser of the fourth column provides the heat for the reboiler of the third column. Similarly, Fig. 6 shows the heat recovery scheme in sequence 2 where heat from the condenser in the third column and from two other intermediate streams is used to provide heat for the reboiler

Table 4. Validation of short-cut model

Component	Ethane		Propane		iso-Butane		n-Butane		Pentanes	
	R*	S*	R	S	R	S	R	S	R	S
Methane	1.36	1.36								
Ethane	95.14	95.09	7.32	7.31						
Propane	3.50	3.53	90.18	89.92	2.00	2.03				
i-Butane					96.00	96.10	4.50	4.61		
n-Butane					2.00	1.87	95.00	94.95		
Butanes			2.50	2.40					3.00	2.82
i-Pentane									33.13	29.58
Pentanes							0.50	0.43		
n-Pentane									23.52	20.93
n-Hexane									26.90	34.62
n-Heptane									13.45	11.97
	R	S								
Hot utility (MW)	89.5	101.7								
Column volume (m^{3})	1,529	1,606								

*R refers reference [23] and S refers results from short-cut model

Table 6. Optimization results of short-cut and rigorous model (sequence 1)

	Short-cut model				Rigorous model			
Column pressure (kPa)	2,003	1,335	587	1,152	2,003	1,335	587	1,152
Reboiler temp. (°C)	87.48	105.05	74.38	143.52	87.12	105.05	74.39	143.52
Condenser temp. (°C)	−10.53	37.74	43.13	86.44	−10.54	37.72	43.13	86.44
Feed stage	8	11	16	14	8	11	16	14
Number of stage	23	32	68	37	23	32	68	37
Reflux ratio	1.04	1.73	10.15	2.10	1.16	1.90	11.55	2.14
Column diameter (m)	6.73	6.38	6.60	5.14	6.40	6.10	6.86	4.72
Column height (m)	−10.35	14.40	30.60	16.65	10.35	14.40	30.60	16.65
Reboiler duty (kW)	2.55×10^4	2.28×10^4	2.86×10^4	1.44×10^4	2.63×10^4	2.40×10^4	3.22×10^4	1.46×10^4
Condenser duty (kW)	1.78×10^4	2.02×10^4	2.83×10^4	1.35×10^4	1.88×10^4	2.14×10^4	3.19×10^4	1.37×10^4
Refrigeration 1 (USD·yr^{-1})	5.01×10^6				5.29×10^6			
CW (USD·yr^{-1})	2.29×10^6				2.48×10^6			
LP (USD·yr^{-1})	5.72×10^6				6.18×10^6			
MP (USD·yr^{-1})	1.53×10^6				1.55×10^6			
Electricity (USD·yr^{-1})	1.77×10^4				1.94×10^4			
Capital cost (USD·yr^{-1})	1.19×10^7				1.19×10^7			
Energy cost (USD·yr^{-1})	1.46×10^7				1.55×10^7			
TAC (USD·yr^{-1})	2.65×10^7				2.74×10^7			

Table 7. Optimization results of short-cut and rigorous model (sequence 2)

	Short-cut model				Rigorous model			
Column pressure (kPa)	2,003	1,447	975	639	2,003	1,447	975	639
Reboiler temp. (°C)	87.48	109.46	134.22	60.52	87.51	109.47	134.25	60.51
Condenser temp. (°C)	−10.53	41.25	73.33	46.47	−10.53	41.30	73.32	46.49
Feed stage	8	11	15	19	8	11	15	19
Number of stage	23	33	36	67	23	33	36	67
Reflux ratio	1.04	1.83	1.48	9.48	1.16	2.03	1.43	10.56
Column diameter (m)	6.73	6.52	5.62	6.03	6.55	6.10	5.18	6.25
Column height (m)	10.35	14.85	15.75	29.25	10.35	14.85	15.75	29.25
Reboiler duty (kW)	2.55×10^4	2.31×10^4	1.83×10^4	2.63×10^4	2.65×10^4	2.45×10^4	1.80×10^4	2.90×10^4
Condenser duty (kW)	1.78×10^4	2.04×10^4	1.72×10^4	2.62×10^4	1.89×10^4	2.18×10^4	1.68×10^4	2.89×10^4
Refrigeration 1 (USD·yr^{-1})	5.01×10^6				5.29×10^6			
CW (USD·yr^{-1})	1.91×10^6				2.10×10^6			
LP (USD·yr^{-1})	4.46×10^6				4.99×10^6			
MP (USD·yr^{-1})	1.94×10^6				1.90×10^6			
Electricity (USD·yr^{-1})								
Capital cost (USD·yr^{-1})	1.15×10^7				1.13×10^7			
Energy cost (USD·yr^{-1})	1.33×10^7				1.43×10^7			
TAC (USD·yr^{-1})	2.46×10^7				2.56×10^7			

in the fourth column. The total heat recoveries obtained using sequences 1 and 2 are 14,634 kW and 22,068 kW, respectively.

For these two selected sequences light components are separated in the first column to minimize the use of high pressure steam and expensive refrigeration. Following optimization and validation using rigorous models in steps two and three the minimum TAC for sequences 1 and 2 are 2.74×10^7 USD·y^{-1} and 2.56×10^7 USD·y^{-1}, respectively. Hence, sequence 2 is selected as the optimal sequence for this case. Sequence 2 separates iso-butane and n-butane in the last column because this separation requires a large number of stages with a high reflux ratio (due to the similar boiling points of the two components).

Figs. 7 and 8 show the hot and cold energy composite curves for sequences 1 and 2 where they are designed at optimal column pressures. These curves are constructed using the temperatures, flow rates and heat capacities of the different hot and cold streams connected to the various condensers and reboilers which are available for heat recovery. Bringing these curves together identifies the pinch point where the maximum heat recovery is achieved using the minimum amount of utilities. For sequence 1 this point is at

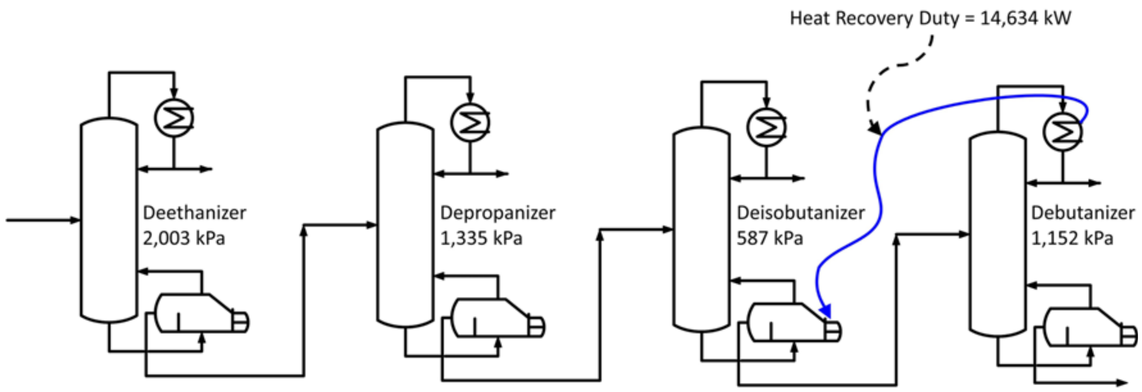

Fig. 5. Heat recovery for the optimized sequence 1 of case A.

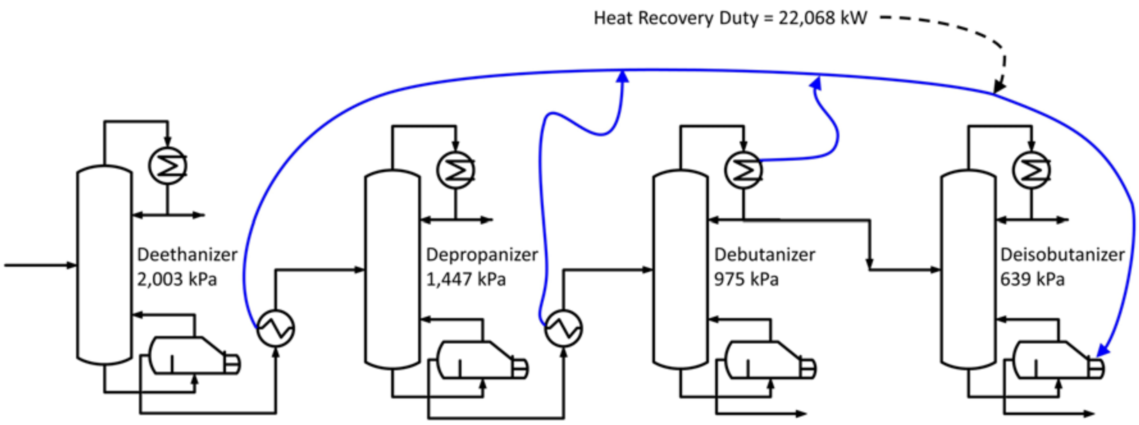

Fig. 6. Heat recovery for the optimized sequence 2 of case A.

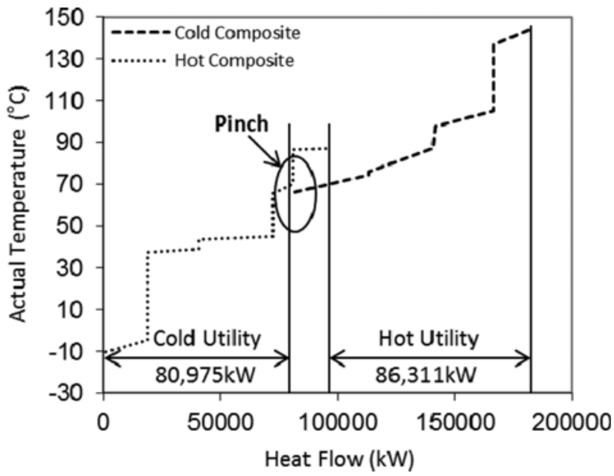

Fig. 7. Energy composite curves for the optimized sequence 1 of case A.

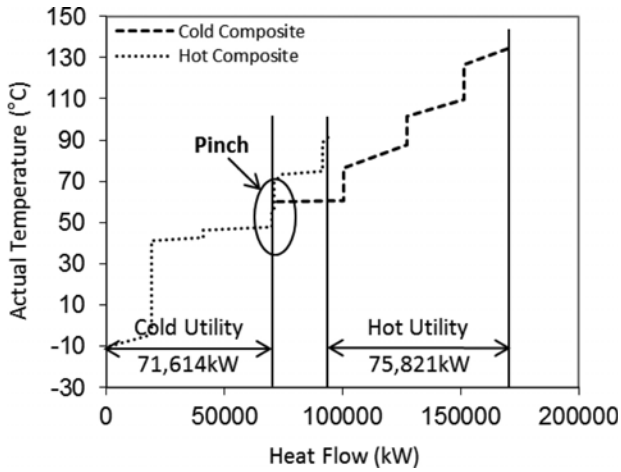

Fig. 8. Energy composite curves for the optimized sequence 2 of case A.

78.19 °C, and this graphical method shows that the minimum hot and cold utilities are 71,723 kW and 54,336 kW, respectively. Similarly, for sequence 2 the pinch is at 77.59 °C, and this gives minimum hot and cold utilities of 65,477 kW and 49,869 kW, respectively.

The difference between the amount of heat recovery possible using the two sequences is due to the different operating temperatures used in the third and fourth columns (a result of the distillation sequencing choices). Sequence 2 is shown to have lower costs compared to the other sequences, and one reason for this is that the arrangement of segeunce can facilitate more heat recovery between columns at the expense of capital investment required for heat exchangers. Also, this case study shows that the short-cut model

Table 8. Results of costing based on short-cut modeling of case B

	Sequence	TAC ($USD\cdot yr^{-1}$)
1	A/BCDE B/CDE C/DE D/E	9.7872×10^6
2	A/BCDE B/CDE CD/E C/D	9.1903×10^6
3	A/BCDE BC/DE B/C D/E	8.5657×10^6
4	A/BCDE BCD/E B/CD C/D	8.3492×10^6
5	A/BCDE BCD/E BC/D B/C	8.1362×10^6
6	AB/CDE C/DE A/B D/E	9.5316×10^6
7	AB/CDE CD/E A/B C/D	8.7707×10^6
8	ABC/DE A/BC B/C D/E	8.4017×10^6
9	ABC/DE AB/C A/B D/E	8.3577×10^6
10	ABCD/E A/BCD B/CD C/D	7.9938×10^6
11	ABCD/E A/BCD BC/D B/C	7.8714×10^6
12	ABCD/E AB/CD A/B C/D	8.0147×10^6
13	ABCD/E ABC/D A/BC B/C	7.9204×10^6
14	ABCD/E ABC/D AB/C A/B	7.7857×10^6

agrees well with results produced from the more rigorous model.

3. Case B

For case B the process under consideration separates a saturated liquid feed of 1,000 $kmol\cdot h^{-1}$ in which four distillation columns (depropanizer, deisobutanizer, debutanizer and depentanizer) are used to produce five products. It is assumed that the distillate recovery is 99% and bottom recovery is 95%. Fourteen different sequences are screened and two of the most promising sequences are selected (sequences 11 and 14) as shown in Table 8. These two sequences are optimized and validated using rigorous simulations as shown in Tables 9 and 10.

In both of these sequences, iso-pentane and n-pentane are split in the first column with a nearly equal split between the molar flow of top and bottom products. Also, separation of iso-butane from n-butane is performed in the last column because of the small difference in relative volatility between these two compounds.

In sequence 11, heat from the feed and from the condenser of the third column is used in two of the reboilers and to heat a number of connecting streams in the sequence. While in sequence 14 heat from the feed and from the condenser of the second column is used in a similar way, providing heat for various reboilers and connecting streams in the sequence (Figs. 9 and 10).

For this case sequence 14 has the best performance in terms of economics considering the set of possible distillation sequences. It can be seen from Fig. 11 and 12 that more heat recovery is possible in sequence 11 compared to sequence 14. However, the economic performance of sequence 14 is optimal for this case because it requires a smaller capital investment.

CONCLUSIONS

A systematic design procedure is presented for the purpose of determining the optimal design of NGL fractionation processes. This design procedure involves the screening of various possible sequences and operating conditions followed by optimization of promising sequences identified through screening. A short-cut model is used to simulate the distillation columns to evaluate the performance of the different sequences for both screening and optimization. In parallel heat integration is incorporated at every step to maximize heat recovery in each of the designs. This is possible by using energy composite curves constructed using the various hot and cold streams available and through analysis which identifies the maximum possible heat recovery (increasing the energy efficiency of the column sequences).

The screening step generates various sequences and sets of operating conditions with "good" and "bad" design solutions in terms of economic performance and energy efficiency. Subsequently the best

Table 9. Optimization results of short-cut and rigorous model (sequence 11)

	Short-cut model				Rigorous model			
Column pressure (kPa)	355	363	696	529	355	363	696	529
Reboiler temp. (°C)	111.06	43.73	98.77	53.56	111.06	43.74	98.81	53.45
Cond. temp. (°C)	35.64	−5.03	59.23	40.80	35.64	−4.11	59.20	40.80
Feed stage	27	4	13	19	27	4	13	19
No. of stage	89	21	31	64	89	21	31	64
Reflux ratio	3.68	4.40	1.39	8.04	4.36	6.66	1.40	8.39
Column diameter (m)	4.51	1.47	2.21	2.45	4.57	1.68	2.29	2.59
Column height (m)	40.05	9.45	13.95	28.8	40.05	9.45	13.95	28.80
Reboiler duty (kW)	1.29×10^4	1.01×10^3	3.05×10^3	4.31×10^3	1.46×10^4	1.47×10^3	3.05×10^3	4.42×10^3
Condenser duty (kW)	1.16×10^4	9.38×10^2	2.97×10^3	4.30×10^3	1.33×10^4	1.40×10^3	2.97×10^3	4.41×10^3
Refrigeration 1 ($USD\cdot yr^{-1}$)	2.61×10^5				3.89×10^5			
CW ($USD\cdot yr^{-1}$)	7.06×10^5				7.69×10^5			
LP ($USD\cdot yr^{-1}$)	1.37×10^6				1.55×10^6			
MP ($USD\cdot yr^{-1}$)								
Electricity ($USD\cdot yr^{-1}$)	2.08×10^3				2.27×10^3			
Capital ($USD\cdot yr^{-1}$)	5.17×10^6				5.43×10^6			
Energy ($USD\cdot yr^{-1}$)	2.34×10^6				2.71×10^6			
TAC ($USD\cdot yr^{-1}$)	7.51×10^6				8.14×10^6			

Table 10. Optimization results of short-cut and rigorous model (sequence 14)

	Short-cut model				Rigorous model			
Column pressure (kPa)	355	593	634	349	355	593	634	349
Reboiler temp. (°C)	111.06	91.55	60.78	25.76	111.06	91.57	60.78	25.76
Cond. temp. (°C)	35.64	45.23	35.89	−6.61	35.64	45.23	35.90	−6.59
Feed stage	27	13	22	6	27	13	22	6
No. of stage	89	30	66	21	89	30	66	21
Reflux ratio	3.68	1.11	6.04	2.12	4.36	1.08	6.70	2.54
Column diameter (m)	4.51	2.23	2.49	0.99	4.57	2.13	2.74	1.07
Column height (m)	40.05	13.50	29.70	9.45	40.05	13.50	29.70	9.45
Reboiler duty (kW)	1.29×10^4	3.22×10^3	4.55×10^3	5.52×10^3	1.46×10^4	3.18×10^3	4.97×10^3	6.25×10^2
Condenser duty (kW)	1.16×10^4	3.11×10^3	4.53×10^3	5.33×10^3	1.33×10^4	3.07×10^3	4.95×10^3	6.06×10^2
Refrigeration 1 (USD·yr^{-1})	1.78×10^5				1.98×10^5			
CW (USD·yr^{-1})	7.46×10^5				8.45×10^5			
LP (USD·yr^{-1})	1.43×10^6				1.65×10^6			
MP (USD·yr^{-1})								
Electricity (USD·yr^{-1})	1.72×10^3				1.88×10^3			
Capital (USD·yr^{-1})	5.22×10^6				5.48×10^6			
Energy (USD·yr^{-1})	2.35×10^6				2.69×10^6			
TAC (USD·yr^{-1})	7.57×10^6				8.17×10^6			

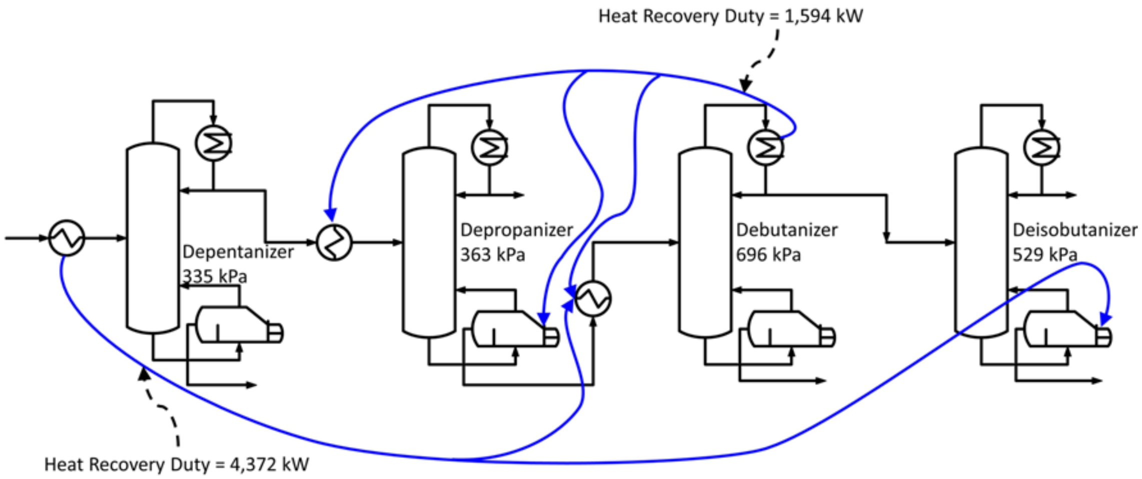

Fig. 9. Heat recovery for the optimized sequence 11 of case B.

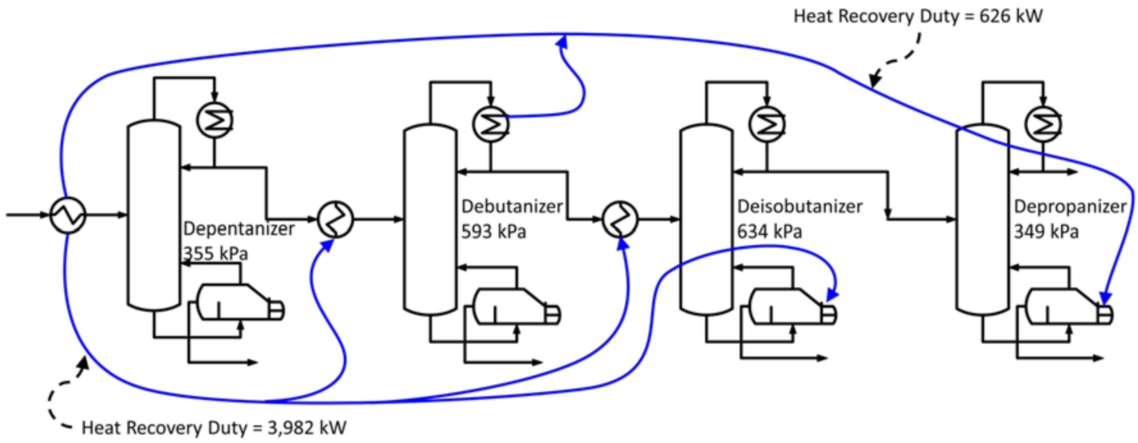

Fig. 10. Heat recovery for the optimized sequence 13 of case B.

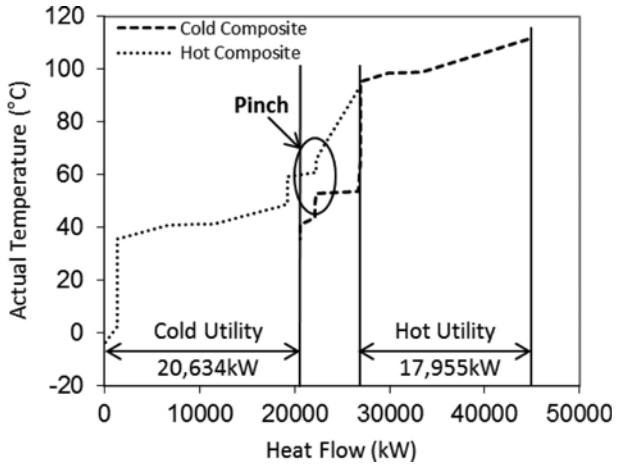

Fig. 11. Energy composite curves for the optimized sequence 11 of case B.

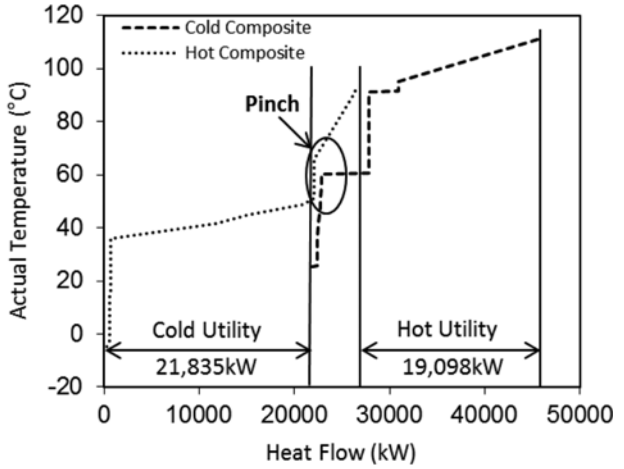

Fig. 12. Energy composite curves for the optimized sequence 13 of case B.

solutions obtained can be optimized in terms of their operating conditions, in particular the column operating pressures. Following these two steps it is recommended to perform rigorous simulations to verify the feasibility of solutions obtained. The main benefits of this design method are its simplicity and the fact that multiple solutions are generated as part of the procedure. In this way the performance of sub-optimal sequences and sequences using sub-optimal conditions are also explored, which can become useful if additional physical constraints are implemented at the facility prohibiting operation at the calculated optimum.

Application of the proposed design methodology to case studies demonstrates the effectiveness of the procedure for the purpose of obtaining economic heat-integrated distillation sequences for NGL fractionation processes. For case studies in question it is shown that the short-cut column model used here gives results similar to those of more rigorous simulations. Hence, this short-cut model is considered appropriate for the proposed design methodology and for the initialization of rigorous simulations which provide some confirmation of the final solutions obtained. The design methodology proposed here can in principle be applied to distillation systems designed for the separation of other multi-component mixtures (other than NGL).

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APPENDIX: ECONOMIC COSTING OF DISTILLATION COLUMNS, HEAT EXCHANGERS AND PUMPS

Operating costs (C_{OP}) are defined as the summation of utility costs used in the distillation columns:

$$C_{OP}=C_{REF}+C_{STM}+C_{CW}+C_E \quad (2)$$

where C_{REF} is the cost of refrigeration (USD·yr^{-1}), C_{STM} is the cost of steam (USD·yr^{-1}), C_{CW} is the cost of cooling water (USD·yr^{-1}) and C_E is the cost of electricity (USD·yr^{-1}). Each of these utility costs is calculated by the unit costs multiplied with amount of utility demand.

The major capital investment required for conventional distillation systems is the purchase costs of columns and auxiliary equipment such as reboiler and condensers. In this study, CEPCI Index (C_{year}^{CEPCI}) is used to update the year-basis of the costs.

Capital Cost (USD):

$$C_{CAP}=(C_{COL}+C_{HEX}+C_{PUMP})(C_{2011}^{CEPCI}/C_{2002}^{CEPCI})F_I \quad (3)$$

Tray Column cost (USD):

$$C_{COL}=N\cdot(839.92\cdot D_C^2+3007.9\cdot D_C-88.368) \quad (4)$$

where F_I is an installation Factor(5.8), the ratio of CEPCI index from 2002 to 2011 is 584.6/395.6, N is the actual number of theoretical stages and D_C is diameter of column (m) [26].

Column diameter can be estimated from Eq. (5), and the largest value would be selected if diameter is not the same throughout the column.

$$D_C=\left(\frac{4\times M\times V}{0.9\times 0.8\times \pi\rho\upsilon}\right)^{0.5} \quad (5)$$

where, V is vapor molar flowrate (kmol·s^{-1}), M is vapor molecular weight (kg·kmol^{-1}), ρ is vapor density (kg·m^{-3}) and υ is vapor flooding velocity (m·s^{-1}).

The cost of a heat exchanger is expressed as [26]:

$$\text{Heat Exchanger cost: } C_{HEX}=N_{US}\left(136.73+\left(118.3\cdot\frac{A_{NET}}{N_{US}}^{0.99}\right)\right)F_P \quad (6)$$

where N_{US} is a total number of units or shells [7], F_P is a pressure adjustment factor (2) [26], A_{NET} is a heat exchange area for vertical heat transfer for the whole network (m^2) [7]:

$$A_{NET}=\Sigma_k^{INTERVALS\,K}\left(\frac{\Delta H_k}{U_k\cdot\Delta T_{LMk}}\right) \quad (7)$$

where K is a total number of enthalpy intervals, ΔH_k is an enthalpy change over interval k (kW), U_k is an overall heat transfer coefficient for interval k (kW·m^{-2}·°C^{-1}), ΔT_{LMk} is a log mean temperature difference for interval k (°C).

Heat exchanger areas are obtained with the assumption that heat transfer coefficients (U) are equal to 1 kW·m^{-2}·°C^{-1} for the use of steam, 0.8 kW·m^{-2}·°C^{-1} for the use of cooling water and 0.65 kW·m^{-2}·°C^{-1} for the use of propane, ethylene, etc. [4].

The cost of a pump is expressed as [26]:

$$\text{Pump cost: } \log(C_{PUMP})=0.6946\cdot\log(W_P)+3.4415 \quad (8)$$

where W_P is the energy required by the pump (kW).

Capital cost is annualized with the following factor:

$$\text{Annualization Factor (yr}^{-1}\text{): } C_{FA}=\frac{i(1+i)^n}{(1+i)^n-1} \quad (9)$$

where, n is the plant life time in years and i is the interest rate per year [26].