Cost Reduction of Fluid Catalytic Cracking Unit in Kaduna Refining and Petrochemical Company using Pinch Technology

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ABSTRACT: Pinch technology is one of the most powerful methodologies of process integration that allows industries to increase their profitability through reductions in energy, water and raw materials consumption. In this study, reduction in the total annual cost of heat exchanger network (HEN) of Fluid Catalytic Cracking (FCC) unit in Kaduna Refining and Petrochemical Company (KRPC), Kaduna was determined. With the help of pinch technology, the reduction was achieved by first determining optimum minimum temperature difference by trading off energy cost and capital cost targets as a function of minimum temperature difference. Thereafter, the total annual cost obtained at the optimum minimum temperature difference was compared with total annual cost of existing design. The results of the analysis showed that the optimum minimum temperature difference was 12℃, the total annual costs of the existing design and the optimum-minimum-temperature-difference based cost were $8.7 and $7.1 Million respectively. This amounted to percentage reduction in the total annual cost of 18.4% which means that about $1.6 Million would been saved annually using the optimum minimum temperature difference to design the HEN of the FCC unit.

I. INTRODUCTION

The severe competition for limited energy in everyday life, technology and development calls for its maximum utilization and this is more so in chemical industries where huge amount of energy is needed. In petroleum and petroleum industries, fluid catalytic cracking (FCC) unit, which is widely regarded as the heart of any refinery, is one of the most energy-consuming units. This energy is usually provided using various utilities like furnace heaters, steam, water and air. However, the consumption of these external utilities can be drastically minimized by using pinch technology to maximize the use of energy available in process streams (Anna, 2011). Pinch technology is the most powerful and widely used process integration methodologies in industries to increase profitability. This can be achieved through reductions in energy (Leni *et al*., 2015; Rathjens *et al*., 2016; Souza *et al*., 2016; Akgun and Ozcelik, 2017; Chang *et al*., 2017; Aguitoni *et al*., 2018; Rathjens and Fieg, 2018).

Similarly, increased profitability can also be achieved through reduction in waste generation, water and raw materials consumption (Sasikala, 2017); and reduction in greenhouse gas emission and environmental pollution control (Li *et al*., 2016; Abdul Aziz *et al*., 2017; Manan *et al*., 2017; Venkatesh, 2019). The technology is based on thermodynamic principles that provides a systematic approach for energy saving with a wide range of applications in many chemical processes (Yoro *et al*., 2019). Although the application of this technology is

most pronounced in oil and gas industries (Marton *et al*., 2017; Akpa *et al*., 2018; Ulyev *et al*., 2018), its use can be found in bioprocessing industry (Anastasovski *et al*., 2107), sugar production industry (Barambu *et al*., 2017), brewing industry (Tibasiima and Okullo, 2017) and cement industry (Verma and Kumar, 2017). It has also been used in addressing project selection problem (Pritam *et al*., 2017).

Typical industrial processes like the FCC units comprise several numbers of hot and cold process streams most of which demand cooling and heating respectively. For this reason, part of the excess heat of the hot process streams is often recovered using heat exchangers to satisfy part of the heat need of the cold process streams. Thereafter, external heaters and coolers are used to achieve the final temperature demand of the process streams. In other words, there must be proper interconnections of hot and cold process streams using heat exchangers to achieve the desire heat recovery and this is where pinch technology comes to play. The most fascinating aspect of the pinch analysis is the ease with which targets can be set even prior to the actual design of the heat exchanger networks. Such targets include heat exchanger area, utility energy (or operating) cost and heat exchanger capital cost targets. All these targets depend on the value of the driving force for the heat flow that is known as the minimum temperature difference (ΔT_{MIN}). Essentially, pinch analysis involves the location of pinch point, which is the temperature level in the process streams where the ΔT_{MIN} is observed (Rokni, 2016).

Several works involving the use of pinch analysis for heat and mass integration of industrial processes have been reported in literatures. Some of the earliest contributors include Linnhoff and Hindmarsh (1983) who used pinch method to design heat exchanger networks, and Gunderson and Naess (1988) who conducted an industrial review on the synthesis of optimal cost heat exchanger network. Other related works include determination of an explicit solution for thermal calculation and synthesis of superstructure heat exchanger networks (Dezhen *et al.*, 2007) and conduction of detailed capital and total cost targets for mass exchanger networks (Hallale and Fraser, 2000). Similarly, Lukman *et al.* (2015) carried out analysis of heat exchanger networks for minimum total annual cost using pinch analysis and Lukman *et al.* (2016) carried out an evaluation of total annual costs of heat exchanger networks using modified pinch analysis.

In this study, the focus was to determine reduction in the total annual cost of the FCC unit of KRPC, Kaduna if pinch technology would have been applied as against the existing design. This was achieved by determining the optimum temperature difference using pinch analysis that guarantees the best trade-off in the energy (or operating) cost and capital cost targets.

II. MATERIALS AND METHODS

A. Materials

The material required for this work was the operating manual of the FCC unit that contains the thermal data. However, these data (as presented in Tables 1 and 2) had already been extracted from the operating manual during the course of our previous work (Olakunle and Abubakar, 2011).

Table 1 presents the thermal data of the existing process streams while Table 2 presents the thermal data of the existing utility streams in the FCC unit. There were nineteen process streams comprising fourteen hot streams and five cold streams. For the utility streams, there were two types of cold utilities: cooling water (CW) and cooling air (CA), and there were also two types of hot utilities: furnace and steam. Four hot process streams were cooled by both water and air, six were cooled by only water, three were cooled by only air and one was not cooled. Similarly, one cold process stream was heated by furnace; another one was heated by steam while the remaining three were not heated.

B. Methods

1.) Estimation of energy (or operating) cost targeting

First, the range of initial ΔT_{MIN} was selected based on the information from literature as it strongly depends on the type of chemical industry under consideration. For petrochemical industries in general, the ΔT_{MIN} is expected not to go beyond 40^0C (Adejoh *et al.*, 2013; Lukman *et al.*, 2016). Hence, the range of initial ΔT_{MIN} was selected to vary from 5 to 40^0C and a step size of 5^0C of this range was adopted. Then, the combined composite curves for each value of the ΔT_{MIN} were constructed using the procedure detailed by Rokni (2016). They are actual temperature – heat load curves and consist of hot (i.e. profiles of heat availability or hot streams in the

Table 1: Thermal data of the process streams in FCC unit (Olakunle and Abubakar, 2011).

Process Stream			Heat Duty (MW)	Actual Temp. (°C)		CP* (MW/°C)
Name	Type	No.		T_S	T_T	
Column Overhead	Hot	H1	19.87	135	36	0.200
Heavy Naphtha	Hot	H2	8.22	169	33	0.060
Light Cycle Oil	Hot	H3	2.21	215	60	0.014
Heavy Cycle Oil Pump Around	Hot	H4	6.66	271	186	0.078
Column Bottom Pump Around	Hot	H5	13.10	350	232	0.111
Column Bottom Product	Hot	H6	1.04	334	80	0.004
Gas Compression Interstage	Hot	H7	1.93	73	36	0.052
Gas Compression Discharge	Hot	H8	3.12	69	36	0.095
Gasoline Product	Hot	H9	6.15	187	40	0.042
Sour Water Stripper Bottom	Hot	H10	1.74	135	40	0.018
C_3/C_4 LPG Product	Hot	H11	0.18	54	40	0.013
Semi Rich Oil	Hot	H12	0.06	41	33	0.008
Debutanizer Overhead	Hot	H13	5.48	67	53	0.391
Sour Water Stripper Overhead	Hot	H14	3.34	126	93	0.101
Fresh Feed	Cold	C1	24.06	70	358	0.084
Mid Stripper Pump Around	Cold	C2	4.82	88	111	0.210
Mid Debutanizer Pump Around	Cold	C3	5.74	105	187	0.070
Debutanizer Feed	Cold	C4	3.74	110	220	0.034
Sour Water	Cold	C5	1.10	53	70	0.065

*CP stands for heat capacity flow rate obtained by dividing heat load by temperature difference

Table 2: Thermal data of the utility streams in FCC unit (Olakunle and Abubakar, 2011)

Process Stream	Heat Duty (MW)		Total (MW)
	Cold Utility		
	Cooling Water (CW)	Cooling Air (CA)	
Column Overhead	1.73	18.14	19.87
Heavy Naphtha	0.61	-	0.61
Light Cycle Oil	-	0.54	0.54
Column Bottom Pump Around	3.46	-	3.46
Column Bottom Product	0.44	0.06	0.50
Gas Compressor Interstage	1.93	-	1.93
Gas Compressor Discharge	1.93	1.19	3.12
Gasoline Product	1.02	3.59	4.61
Sour Water Stripper Bottom	0.64	-	0.64
C3/C4 LPG Product	0.18	-	0.18
Semi Rich Oil	0.06	-	0.06
Debutanizer Overhead	-	5.48	5.48
Sour Water Stripper Overhead	-	3.34	3.34
Total	**12.00**	**32.88**	**44.88**
	Hot Utility		
	Furnace	Steam	
Fresh Feed	9.04	-	9.04
Debutanizer Overhead	-	2.20	2.20
Total	**9.04**	**2.20**	**11.24**

the process) and cold (i.e. heat demand or cold streams in the process) composite curves. The construction of each curve was carried out by adopting 'Problem Table Method'. In this method, the respective streams' actual temperatures (both

source and target) were ranked either in decreasing or increasing order. Then CPs of all streams that fell into a particular interval of the ranked temperatures were added and the sum were divided by the temperature difference in that interval to obtain the heat load for that interval. For instance, the heat load (ΔH_n) for nth interval of temperature difference (ΔT_n) was determined by

$$\Delta H_n = (\textstyle\sum CPs) \times (\Delta T_n) \qquad (1)$$

With the lowest temperature assigned either a zero or an arbitrary heat load, the calculated heat loads were then cumulated over the assigned heat load so that every temperature corresponded to a particular heat load. Finally, the temperatures were then plotted against heat loads to give the composite curves. Now, the combined (i.e. both hot and cold) composite curves for a particular ΔT_{MIN} produced three distinct regions (Fig. 1). These regions were (1) an overlap region that represented the maximum possible scope of heat recovery within the process, (2) an overshoot of the hot composite curve that represented the minimum cold utility requirement ($Q_{C,MIN}$) and (3) an overshoot of the cold composite curve that represented the minimum hot utility requirement ($Q_{H,MIN}$).
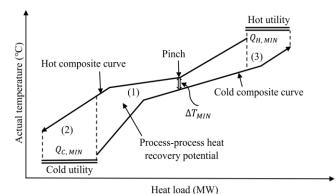

Fig. 1: Combined composite curves

The combined composite curves also gave valuable insights into the selection of appropriate utilities that would provide these heat requirements. The first condition for selection is that the hot utilities must have higher temperatures than the temperatures in the region of the cold composite curve where it will provide heat. Conversely, the cold utilities must have lower temperatures than the temperatures in the region of the hot composite curve where it will receive heat. In conformity with this condition, there were suggested, based on experiences, ranges of the ΔT_{MIN} for petrochemical industries which should be maintained between the process streams and utility streams (Auta *et al.*, 2012), and this informed the choice of the temperature range for each utility in Table 3. The second condition is that utility with lower cost is preferable if more than one utility meet the first condition. The choice of annual

unit costs was based on information obtained from literatures (Turton *et al.*, 2009; Towler and Sinnott, 2013).

Table 3: Selected utilities (Turton *et al.*, 2009; Auta *et al.*, 2012; Towler and Sinnott, 2013).

Utility	Type	Temperature (°C)		Unit cost, C_U ($/MW, year)
		Source (T_S)	Target (T_T)	
Water (CW)	Cold	20	45	11,000
Air (CA)	Cold	20	80	5,500
Fuel	Hot	350	400	350,000

The grand composite curves were the next to be constructed and the 'Problem Table Method' was also adopted. However, the grand composite curves are different from the combined composite curves in the following senses:

- Both the cold and hot process streams' temperatures were combined together to obtain only one plot.
- Instead of the actual temperatures, the interval temperatures as obtained from Equations (2) were used.
- After obtaining the heat load in each interval using Equation (1), the surplus heat load from one interval to the next down the column of each interval was cascaded and just enough heat was introduced to the top of the cascade to eliminate all the negative values.
- The interval temperatures were then plotted against the cascaded heat loads to give the grand composite curve.

$$\text{Hot streams: } T_{Int} = T_{Act} - \Delta T_{MIN}/2 \qquad (2a)$$

$$\text{Cold streams: } T_{Int} = T_{Act} + \Delta T_{MIN}/2 \qquad (2b)$$

where T_{Act} and T_{Int} are actual and interval temperatures respectively

The grand composite curves, if correctly constructed would also give the same values of $Q_{C,MIN}$ and $Q_{H,MIN}$ as those obtained from combined composite curves. Next, the pinch point or pinch temperature (T_P) was located while the hot stream pinch ($T_{P,hot}$) and cold stream pinch ($T_{P,cold}$) temperatures were determined. Finally, using the selected utilities, the energy cost targeting for each value of the selected ΔT_{MIN} was estimated using the following equation (Paiko *et al.*, 2017; Yoro *et al.*, 2018).

$$C_{ENERGY}(\$) = \textstyle\sum Q_U \cdot C_U \qquad (3)$$

where Q_U = Heat duty of an utility, MW and C_U = Unit cost of an utility, $/MW, year.

2.) Estimation of HEN capital cost targeting

The capital cost is the heat exchangers' cost, which is a function of the sizes or areas of the heat exchangers. Therefore, the minimum area (A_{MIN}) targeting was first estimated using the combined composite curves with the utility streams included. It was assumed that the area was evenly distributed between the units since the actual area distribution could not be determined ahead of design. Then the combined composite curves were divided into a set of adjoining sections such that within each section, the hot and cold composite curves did not

change slope (Fig. 2). This allowed the hot streams in any section, at any point, to exchange heat with the cold streams at the temperature vertically below it (i.e. each section represented one heat exchanger). Here, the heat exchange was assumed to be "vertical" (pure counter-current heat exchange).
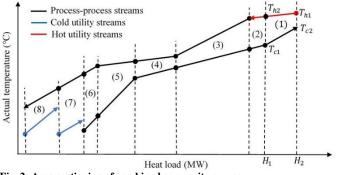

Fig. 2: Area sectioning of combined composite curves.

Therefore, the surface area for each section (A_i) was calculated using the following widely accepted heat exchanger equation defined in Rathjens and Fieg (2019).

$$A_i = \Delta H_i / (U_i \times \Delta T_{LM-i}) \tag{4}$$

where ΔH_i = heat load, U_i = Overall heat transfer coefficient as presented in Table 4 (Towler and Sinnott, 2013) and ΔT_{LM-i} = Log mean temperature difference. Subscript i stands for a particular section.

For example, in section 1 in Figure 2, the heat load (ΔH_1) and log mean temperature difference (ΔT_{LM-1}) were obtained as follows:

$$\Delta H_1 = H_2 - H_1 \tag{5}$$

$$\Delta T_{LM-1} = \frac{(T_{h1} - T_{c2}) - (T_{h2} - T_{c1})}{\ln[(T_{h1} - T_{c2})/(T_{h2} - T_{c1})]} \tag{6}$$

Table 4: Selected overall heat transfer coefficients (Towler and Sinnott, 2013).

Heat exchanging streams	Overall heat transfer coefficient, U (MW/m² °C)
Process stream – Process stream	250
Process stream – Fuel	100
Process stream – Cooling water	500
Process stream – Cooling air	625

The total area of the HEN (A_{MIN}) above and below the pinch became equal to the sum of the areas calculated above and below the pinch respectively. That is;

$$A_{MIN} = \sum A_i \tag{7}$$

It should be noted that the actual A_{MIN} required is generally within 10% of the area target as calculated using this procedure according to Smith (2005).

Next, the minimum number of units (N_{MIN}) targeting was separately estimated for sections above and below the pinch. The following equation (Rokni, 2016; Ateeq *et al.*, 2017) was used for each section.

$$N_{MIN} = N_H + N_C + N_U - 1 \tag{8}$$

where N_H = Number of hot streams, N_C = Number of cold streams and N_U = Number of utility streams.

The numbers of hot and cold streams were determined using the $T_{P,hot}$ and $T_{P,cold}$ respectively to divide the process streams into two regions (i.e. above the pinch and below the pinch). On the other hand, the number of utility streams depends not only the number of process streams that required the utilities but also on the thermal constraints of the utilities.

Finally, the HEN capital cost (C_{HEN}) targeting was estimated using the following equation (Paiko *et al.*, 2017; Yoro *et al.*, 2018).

$$C_{HEN}(\$) = [N_{MIN}\{a + b(A_{MIN}/N_{MIN})^c\}]_{AP} + [N_{MIN}\{a + b(A_{MIN}/N_{MIN})^c\}]_{BP} \tag{9}$$

Equation (9) was derived from the following well-established capital cost of a single heat exchanger equation (Hojjati *et al.*, 2004; Skolpap and Owat, 2018).

$$\text{Heat Exchanger Cost } (\$) = a + b(Area)^c \tag{10}$$

where a, b and c are known as cost law constants whose values depend on materials of construction among other factors. For a carbon steel shell and tube exchanger which was adopted in this study, the values of these constants are a = 16,000, b = 3,200 and c = 0.7 (Linhoff and Vredeveld, 1984; Hojjati *et al.,* 2004). Subscripts AP and BP stand for above the pinch and below the pinch.

3.) Determination of reduction in the total annual cost

The energy cost targeting (C_{ENERGY}) calculated in section 2.2.1 and the capital cost targeting (C_{HEN}) calculated in section 2.2.2 were summed to give the total cost targeting (C_{TOTAL}). These three costs (i.e. energy, capital and total costs) were then plotted against their respective minimum temperature differences (ΔT_{MIN}) on the same graph (presented as Figure 5 in result section). It should be noted that in plotting these data, a regression analysis was used so that lines of best fit were obtained. Because C_{ENERGY} increases with increase in ΔT_{MIN} while C_{HEN} decreases with increase in ΔT_{MIN}, the C_{TOTAL} produces a minimum. The process of determining this minimum is referred to as trade-off between energy cost and capital cost in economic analysis of chemical plants. The

minimum C_{TOTAL} is known as optimum annual total cost and the minimum temperature difference at which this optimum C_{TOTAL} is obtained is known as optimum minimum temperature difference (Attarakih *et al.*, 2013; Rokni, 2016; Tibasiima and Okullo, 2017). Next is the determination of existing annual energy cost (C^o_{ENERGY}) of the FCC unit which was achieved by applying Equation (3) to the existing utility thermal data. Using the calculated existing C^o_{ENERGY}, the existing total annual cost (C^o_{TOTAL}) was traced out from the same Figure 5. Finally, the percentage reduction in the total annual cost (% C^R_{TOTAL}) of the FCC unit was determined using Equation (11).

$$\% \, C^R_{TOTAL} = \frac{C^o_{TOTAL} - C_{TOTAL}}{C^o_{TOTAL}} \times 100 \qquad (11)$$

III. RESULTS AND DISCUSSION

A. Presentation of Results

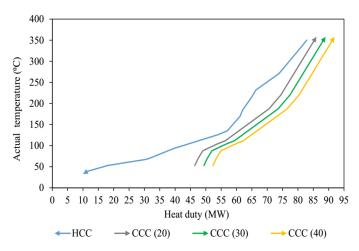The results obtained from this work are presented in Figs. 3 to 7 and Tables 5 to 6.

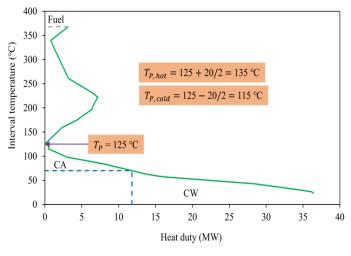Fig. 3: Combined composite curves for ΔT_{MIN} of 20, 30 and 40℃.
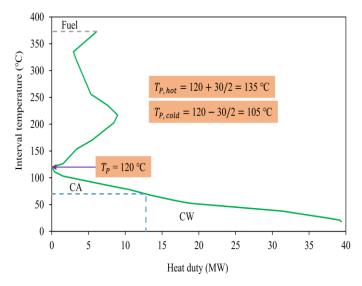

Fig. 4: Grand composite curve for ΔT_{MIN} of 20℃.
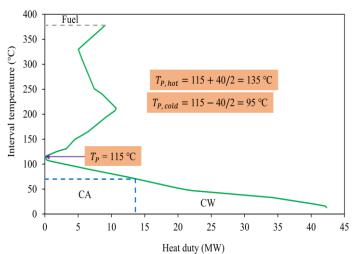

Fig. 5: Grand composite curve for ΔT_{MIN} of 30℃.

Fig. 6: Grand composite curve for ΔT_{MIN} of 40℃.

Table 5: Cost targeting at different ΔT_{MIN}.

ΔT_{MIN} (°C)	A_{MIN} (m²)		N_{MIN}		C_{ENERGY} (Million $/Yr.)	C_{HEN} (Million $/Yr.)
	AP	BP	AP	BP		
5	676	9827	1	36	0.66	6.74
10	678	9023	1	36	0.80	6.40
15	3770	4577	9	33	1.17	5.98
20	3302	4147	9	33	1.38	5.58
25	2688	3776	12	32	1.96	5.29
30	2765	3361	10	32	2.53	4.97
35	2643	3283	10	32	3.04	4.88
40	2594	2924	10	32	3.58	4.65

Table 6: Annual energy cost of existing HEN.

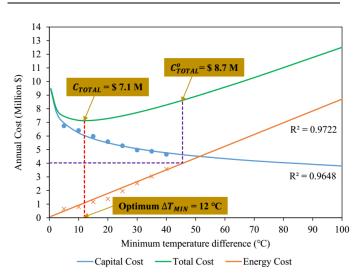| Utility	H_U (MW)	C_U ($/MW, year)	Cost (Million $/year)
CW	12.00	11000	0.132
CA	32.88	5500	0.181
Steam	2.20	250000	0.550
Fuel	9.04	350000	3.164
C^o_{ENERGY}			4.027

Fig. 7: Energy cost and capital cost trade-off.

B. Discussion of Results

1.) Combined composite curves

Hot composite curve (HCC) and some cold composite curves (CCCs) are plotted together as the combined composite curves in Fig. 3. It should be noted that although ΔT_{MIN} of 5, 10 ,15, 20, 25, 30, 35 and 40°C were used for the analysis, only the composite curves for ΔT_{MIN} of 20, 30 and 40°C were presented here for the sake of clarity as they all follow similar trends. The essence of these curves is to find the minimum hot and cold utility requirement. From these curves therefore, the cold utility requirement (the difference between the lowest heat load of the hot composite curve and the lowest heat load of the cold composite curve) was much higher than the hot utility requirement (the difference between the highest heat load of the hot composite curve and the highest heat load of the cold composite curve). Secondly, theses curves show that the higher the ΔT_{MIN}, the higher the hot and cold utility requirements. This is in agreement with what has been reported in the literatures (Smith, 2005; El-Halwagi, 2006).

2.) Grand composite curves

The grand composite curves for ΔT_{MIN} of 20, 30 and 40°C are presented in Figs. 4 to 6. The reason given in section 3.2.1 regarding the selected values of ΔT_{MIN} for the composite curves is equally applicable here. These curves show the pinch point, which increased with increase in the ΔT_{MIN}. Again, they also show the hot and the cold pinch points: the hot pinch point remained unchanged while the cold pinch point decreased with increase in the ΔT_{MIN}. These trends are not universal, rather they depend on the nature of the heat exchanger network under

consideration. Finally, the cold utility levels shown in these curves were done in such a manner that the cooling water served the coldest portion of the curves up to 70°C while the cooling air served the remaining portion. The reason is to make sure that the discharged warm water does not have significant negative impact on the environment (i.e. land or river).

3.) Cost targeting

Table 5 presents the minimum areas and numbers of units targeting, calculated energy costs, and HEN capital costs for all the initially selected ΔT_{MIN}. It is obvious to see that the total area targeting (i.e. sum of areas above and below the pinch) and the total number of units targeting (i.e. sum of numbers of units above and below the pinch) depended on the ΔT_{MIN}, with the former showing very much stronger dependence. Specifically, the total area targeting decreased with increase in the ΔT_{MIN}. This is expected because the larger the ΔT_{MIN}, the less the amount of heat that can be exchanged between cold and hot process streams. In contrast, the total minimum number of unit targeting showed slight increase from 37 at ΔT_{MIN} of 5 and 10°C to 42 at higher ΔT_{MIN}.

Regarding the cost targeting, the trend in the total area targeting translated to having higher hot and cold utility requirements which was why the energy cost targeting also increased with the increase in the ΔT_{MIN}. It should be noted that the energy cost depends only on the quantity and type of utilities based on Equation (3) and not the number of heat exchangers or units. On the other hand, the HEN cost is proportional to both the area and the number of units based on Equation (9). Therefore, since the trend of the total area targeting with respect to the ΔT_{MIN} was in contrast with that of the total minimum number of unit targeting, then the overall trend of HEN capital cost targeting with respect to the ΔT_{MIN} would depend on the relative dominance of the area and the number of units. As can be seen in Table 5, the HEN capital cost targeting decreased with increase in the ΔT_{MIN} showing that the effect of the area on HEN cost outweighed that of the number of units.

4.) Reduction in total annual cost

The annual energy cost of the existing HEN was calculated, and the result is presented in Table 6. Using the energy cost and capital cost trade-off diagram presented in Figure 7, the existing total annual cost was found to be $8.7 Million. The total annual cost targeting ($7.1 Million) is the minimum total annual cost in Fig. 7 and the temperature (12°C) at which this is obtained is called the optimum minimum temperature difference (optimum ΔT_{MIN}). Using Equation (11), the percentage reduction in the total annual cost was obtained to be 18.4%. This means that if the design of the HEN for FCC unit was done using the optimum ΔT_{MIN}, 18.4% of the total cost, which translated to about $1.6 Million, would have been be saved annually.

IV. CONCLUSION

Pinch technology was used to determine the reduction in the total annual cost of HEN of FCC unit and the following conclusions were drawn from the results.

a) The total annual cost of the existing design of the FCC unit was found to $8.7 Million.

b) The optimum minimum temperature difference of the FCC unit was found to be 12℃.

c) The total annual cost if the optimum minimum temperature difference was used to design the FCC unit was found to be $7.1 Million.

d) The percentage reduction in the total annual cost using optimum minimum temperature difference was found to be 18.4%.

V. ACKNOWLEDGEMENT

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