Application of Pinch Analysis for Energy Efficient Design of Distillation Column in Petroleum Refinery

By

Kaleem Ullah

(Registration No: 00000329117)

Department of Chemical Engineering

School of Chemical and Materials Engineering

National University of Sciences & Technology (NUST)

Islamabad, Pakistan

(2024)

Application of Pinch Analysis for Energy Efficient Design of Distillation Column in Petroleum Refinery

By

Kaleem Ullah

(Registration No: 00000329117)

A thesis submitted to the National University of Sciences and Technology, Islamabad,

in partial fulfillment of the requirements for the degree of

Master of Science in
Process Systems Engineering

Supervisor: Dr. Muhammad Ahsan

Co-Supervisor: Dr. Iftikhar Ahmed

School of Chemical and Materials Engineering

National University of Sciences & Technology (NUST)

Islamabad, Pakistan

(2024)

THESIS ACCEPTANCE CERTIFICATE

Certified that final copy of MS thesis written by Mr Kaleem Ullah (Registration No 00000329117), of School of Chemical & Materials Engineering (SCME) has been vetted by undersigned, found complete in all respects as per NUST Statues/Regulations, is free of plagiarism, errors, and mistakes and is accepted as partial fulfillment for award of MS degree. It is further certified that necessary amendments as pointed out by GEC members of the scholar have also been incorporated in the said thesis.

Signature: ____________________

Name of Supervisor: Dr Muhammad Ahsan

Date: 16-09-2024

Signature (HOD): ____________________

Date: 16-09-2024

Signature (Dean/Principal): ____________________

Date: 18/9/24

Form TH-1

National University of Sciences & Technology (NUST)

MASTER'S THESIS WORK

Formulation of Guidance and Examination Committee (GEC)

ame: Kaleem Ullah NUST Reg No: 00000329117

epartment: Department of Chemical Engineering Specialization: Master of Science in Process System Engineering

'edit Hour Completed: 21.0 CGPA: 3.93

Course Work Completed

S/No:	Code:	Title:	Core/Elective:	CH:	Grade:
1.	RM-898	Research Methodology	Additional	2.0	Q
2.	PSE-801	Process Systems Theory	Compulsory	3.0	A
3.	TEE-820	Process Intensification	Compulsory	3.0	A
4.	PSE-852	Process Modelling and Simulation	Compulsory	3.0	B+
5.	PSE-802	Optimization and Decision Analysis	Compulsory	3.0	A
6.	PSE-823	Advanced Process Dynamics and Control	Compulsory	3.0	A
7.	EME-902	Numerical Methods In Chemical Engineering	Elective	3.0	A
8.	CSE-801	Computational Fluid Dynamics	Elective	3.0	A

te 12 - Oct - 2022 Student's Signature

Thesis Committee

Name: Muhammad Ahsan (Supervisor) Signature
Department: Department of Chemical Engineering

Name: Iftikhar Ahmad (Cosupervisor) Signature
Department: Department of Chemical Engineering

Name: Umair Sikander (Internal) Signature
Department: Department of Chemical Engineering

Name: Muhammad Nouman Aslam Khan (Internal) Signature
Department: Department of Chemical Engineering

e: 12 - Oct - 2022 Signature of Head of Department:

APPROVAL

e: 12 - Oct - 2022 Signature of Dean/Principal:

School of Chemical & Materials Engineering (SCME) (SCME) H-12 Campus, Form: TH-04

National University of Sciences &Technology (NUST)

MASTER'S THESIS WORK

We hereby recommend that the dissertation prepared under our supervision by

Regn No & Name: 00000329117 Kaleem Ullah

Title: Application of Pinch Analysis for Energy Efficient Design of Distillation Column in Petroleum Refinery.

Presented on: 29 Aug 2024 at: 1430 hrs in SCME (Seminar Hall)

Be accepted in partial fulfillment of the requirements for the award of Master of Science degree in Process Systems Engineering.

Guidance & Examination Committee Members

Name: Dr Umair Sikander Signature:

Name: Dr Muhammad Nouman Aslam Khan Signature: Nouman

Name: Dr Iftikhar Ahmad (Co-Supervisor) Signature:

Supervisor's Name: Dr Muhammad Ahsan Signature:

Dated: 10-09-2024

Head of Department
Date 10-09-24

Dean/Principal
Date 10/9/24

School of Chemical &Materials Engineering (SCME)

AUTHOR’S DECLARATION

I, Kaleem Ullah, hereby state that my MS thesis titled “Application of Pinch Analysis for Energy Efficient Design of Distillation Column in Petroleum Refinery” is my own work and has not been submitted previously by me for taking any degree from National University of Sciences and Technology, Islamabad or anywhere else in the country/ world.

At any time if my statement is found to be incorrect even after I graduate, the university has the right to withdraw my MS degree.

Name of Student: Kaleem Ullah

Date: 29th August, 2024

Plagiarism Undertaking

I solemnly declare that the research work presented in the thesis titled "**Application of pinch analysis for the efficient design of distillation column in petroleum refinery**" is solely my research work with no significant contribution from any other person. Small contribution/ help wherever taken has been duly acknowledged and that complete thesis has been written by me. -

I understand the zero-tolerance policy of the HEC and National University of Sciences and Technology (NUST), Islamabad towards plagiarism. Therefore, I as an author of the above titled thesis declare that no portion of my thesis has been plagiarized and any material used as reference is properly referred/cited.

I undertake that if I am found guilty of any formal plagiarism in the above titled thesis even after award of MS degree, the University reserves the rights to withdraw/revoke my MS degree and that HEC and NUST, Islamabad has the right to publish my name on the HEC/University website on which names of students are placed who submitted plagiarized thesis.

Student Signature: ____________

Kaleem Ullah

DEDICATION

To my Loving, Caring and Supportive Family

ACKNOWLEDGEMENTS

First and foremost, praise is to Almighty Allah, Who bestowed upon me with His blessings and guided me in completing this task. It would not have been possible without the support of my family and the foremost help and support of several individuals/colleagues who were available all the time and helped me with their suggestions to compile this work.

I extend my heartfelt gratitude to my supervisor, **Dr. Muhammad Ahsan**, for his timely advice and valuable, practical approach during this research work. It would be worth mentioning here that it is only his constant pursuit that leads to completing my work. There were several hurdles and difficulties during my research work, but he remained not always available but chased me to complete the work more efficiently.

I also express my appreciation to **Dr. Iftikhar Ahmad, Dr. Umair Sikander and Dr. Muhammad Nouman Aslam Khan**, members of my Guidance and Examination Committee (GEC), for their invaluable assistance. Furthermore, the contributions of numerous postgraduate students are acknowledged. They have been an incredible source of encouragement and support.

Kaleem Ullah

TABLE OF CONTENTS

ACKNOWLEDGEMENTS **IX**

TABLE OF CONTENTS **X**

LIST OF TABLES **XII**

LIST OF FIGURES **XIII**

LIST OF SYMBOLS, ABBREVIATIONS AND ACRONYMS **XIV**

ABSTRACT **XV**

CHAPTER 1: INTRODUCTION **1**
1.1 Background and Motivation **1**
1.2 Objectives **4**
1.3 Organization of the Thesis **5**

CHAPTER 2: LITERATURE REVIEW **6**

CHAPTER 3: METHODOLOGY **12**
3.1 Process description **12**
3.2 Methodology **14**
3.3 Process Simulation **15**
3.3.1 Simulation Validation 16
3.4 Exergy Analysis Formulations **17**
3.4.1. Exergy Analysis of Heat Exchangers 17
3.4.2. Exergy Analysis of Heat Exchanger Network 18
3.4.3. Exergy Analysis of Preflash column 19
3.4.4. Exergy Analysis of Furnace 20
3.4.5. Exergy Analysis of crude distillation column 20
3.5 Pinch Analysis **21**
3.5.1. Pinch analysis sequence 21
3.5.2. Process & utility streams data 22
3.5.3. Targets 25
3.5.4. Energy-Efficient Design of the Distillation Column 25
3.5.5. Energy Optimization of the Distillation Column 26

CHAPTER 4: RESULTS AND DISCUSSION **27**
4.1 Exergy Analysis **27**
4.1.1. Exergy Destruction 30
4.1.2. Exergy Efficiency 33
4.1.3. Exergy Destruction and Efficiency of Other Process Units 35
4.1.4. Effect of number of trays on overall exergy efficiency and irreversibility 37
4.2 Pinch Analysis **38**

4.2.1. Composite Curve 38
4.2.2. Grand Composite Curve 39
4.2.3. Heat Exchanger Network (HEN) design – base case 40
4.2.4. Base case performance 41
4.2.5. Heat exchangers & utilities data – base case 41
4.2.6. Heat Exchanger Network (HEN) design – optimized case 43
4.2.7. Optimized case performance 43
4.2.8. Heat exchangers & utilities data – optimized case 44
4.2.9. Comparison of the base case and optimized designs 45
4.3 Economic Analysis 46
4.3.1. Total Direct and indirect cost 47
4.3.2. Production cost 51
4.3.3. Profitability Analysis 53
4.3.4. Depreciation 54
4.3.5. Payback period calculation 55
4.3.6. Comparative analysis of base and optimized case based on economic analysis 56
4.3.7. Discussion of the results 57
4.4 Limitations of the study 59

CHAPTER 5: CONCLUSION 60
5.1 Conclusion 60
5.2 Future Work 60

REFERENCES 62

LIST OF TABLES

Page No.

Table 1: Comparison between HYSYS results and actual crude distillation data 16
Table 2: Process Hot and Cold Stream Data for Heat Network Design 23
Table 3: Operating Conditions and exergy data of process streams 27
Table 4: Exergy destruction of all the individual heat exchangers 31
Table 5: Exergy efficiencies of individual heat exchangers 33
Table 6: Exergy efficiency and exergy destruction of all process units 35
Table 7: Base case heat exchangers 42
Table 8: Base case utilities.. 43
Table 9: Optimized case heat exchangers 44
Table 10: Optimized case utilities... 45
Table 11: Cost comparison of base and optimized case based on economic.................... 46
Table 12: Cost of Raw Material and Product... 47
Table 13: Equipment cost summary from Aspen Economic Analyzer............................. 48
Table 14: Total Direct Cost Estimation 50
Table 15: Total Indirect Cost Evaluation.. 51
Table 16: Direct and Annual Production Cost.. 52
Table 17: A comparative analysis based on economic analysis between the base case and optimized case... 56

LIST OF FIGURES

Page No.

Figure 1: Enhancement potential for energy conservation in petroleum refineries [4] 2
Figure 2: Flowsheet of a crude distillation unit 13
Figure 3: HEN of crude distillation unit 13
Figure 4: General phases of the study 14
Figure 5: Flowsheet of crude distillation unit in HYSYS 15
Figure 6: Crude preheat train (1 & 2) of crude distillation unit 15
Figure 7: Overall HYSYS model of the distillation column and side strippers 16
Figure 8: General exergy balance of the system 21
Figure 9: Pinch Analysis Steps 22
Figure 10: Grid diagram displaying the pinch segments above and below the pinch point 24
Figure 11: Process streams data in the AEA environment 24
Figure 12: Utility streams data in the AEA environment 25
Figure 13: Target values of heat exchanger network 25
Figure 14: Base case heat exchanger network. 26
Figure 15: Optimized heat exchanger network (HEN) 26
Figure 16: Contribution of heat exchangers to the overall exergy destruction of the entire HEN 33
Figure 17: Exergy efficiencies of all the heat exchangers of HEN 35
Figure 18: Effect of number of trays on overall exergy efficiency and irreversibility 38
Figure 19: Composite curve of heat exchanger network 39
Figure 20: Grand Composite Curve of heat exchanger network 40
Figure 21: maximum energy recovery base case network 41
Figure 22: Performance data of base case network 41
Figure 23: Optimized and final recovery network 43
Figure 24: Performance data of optimized case network 44
Figure 25: crude oil prices and prices of six refined products 47
Figure 26: Net Present Value versus Plant life for the process 56

LIST OF SYMBOLS, ABBREVIATIONS AND ACRONYMS

CDU	Crude Distillation Unit
PA_Draw	Pumparound draw-off
PA	Pinch Analysis
AEA	Aspen energy analyzer
APEA	Aspen Process Economic Analyzer
HEN	Heat Exchanger Network
HS	Hot Streams
CS	Cold Streams
DCU	Delayed Cocking Unit
E	Exergy
EPH	Physical exergy
EPT	Potential exergy
EKN	Kinetic exergy
ECH	Chemical exergy
I	Irreversibility
Iex	Irreversibility of heat exchanger
IHEN	Irreversibility of heat exchanger network
Ed	Exergy destruction
EH,in	Exergy of the hot inlet stream
EH,out	Exergy of the hot outlet stream
EC,in	Exergy of the cold inlet stream
EC,out	Exergy of the cold outlet stream
T	Temperature
T_o	Reference temperature
S_{gen}	Entropy generated
η_{ex}	Exergy efficiency of heat exchanger
η_{HEN}	Exergy efficiency of heat exchanger network
E_p	Exergy produced

ABSTRACT

This study focuses on optimizing the crude preheat train within a petroleum refinery's crude distillation unit (CDU) through the integration of Exergy Analysis, Pinch Analysis (PA), and Economic Analysis for the developed model in Aspen HYSYS. Exergy Analysis is employed to identify energy inefficiencies within the CDU. Initially, the steady-state exergy analysis was carried out using an Aspen HYSYS model to quantify the exergy destructions and exergy efficiencies of all the individual heat exchangers and the overall Heat Exchanger Network. The overall exergy destruction and exergy efficiency of the HEN was 17611.21 kW and 63.34%, respectively. Exergy Analysis reveals that the energy efficiency can be improved for some of the heat exchangers in the network as they contribute 56 % to the overall exergy destruction of HEN. PA, facilitated by Aspen Energy Analyzer (AEA), for pinch analysis of the existing preheat trains of crude distillation unit. The optimized case is also developed for more energy integrated and efficient design of crude preheat train and distillation unit. The optimized case achieved the same energy requirements as of base case with two fewer heat exchangers, reducing the overall capital cost of the crude distillation unit by $ 95716.08 despite the increase in heating and cooling utility costs. Finally, a payback period of 1.18 years has been calculated through an economic analysis for being a crude distillation unit operational.

Keywords: Pinch Analysis, Heat Integration, Exergy efficiency, Exergy destruction, Economic Analysis, Heat Conservation

CHAPTER 1: INTRODUCTION

1.1 Background and Motivation

The importance of improved energy efficiency in industrial production is growing due to its environmental, economic, and commercial consequences. Enhanced energy efficiency not only provides direct economic benefits, such as increased competitiveness and higher production, but also shows great promise in reducing CO_2 emissions resulting from the use of fossil fuels. The emphasis on energy efficiency is especially important for several enterprises that operate in well-established and energy-intensive sectors. These industries frequently encounter energy prices that surpass 30% of their total production expenses, requiring a strategic emphasis on energy efficiency to sustain competitiveness in the future [1]. It is generally acknowledged that the petroleum refinery is one of the industries that generates the most energy-intensive output. As a result, it is of the utmost importance to establish an energy-efficient approach to the design as well as the operation of processes. Due to the fact that a significant amount of heat is required to fractionate the oil, the crude oil processing that takes place in the crude distillation unit (CDU) consumes the majority of the energy. The CDU is responsible for approximately 15 to 25 percent of the total energy consumed throughout the refining process [2, 3]. Figure 1 depicts numerous options for energy savings in the setting of a petroleum refinery. As shown in Figure 1, the crude distillation unit offers a large possibility for energy savings. As a result, improving the energy efficiency of any process is always a top priority for ensuring its profitability and sustainability.

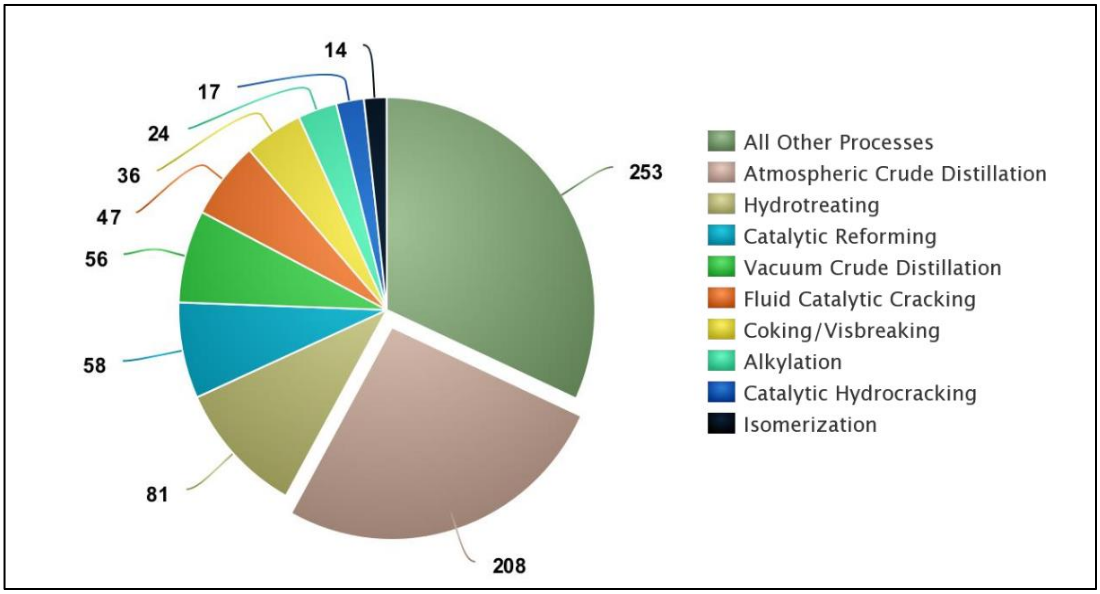


Figure 1: Enhancement potential for energy conservation in petroleum refineries [4]

Getting end products such as kerosene, gasoline, and other important fuels from crude oil requires using more than one unit. Initial treatment involves putting the crude oil into an atmospheric distillation column at a very high temperature (almost 360 °C). This temperature can change depending on the processing conditions and the source of the crude oil. The crude oil is heated up first in a pre-heat train. It then goes into the furnace, which uses the most energy. Improving the energy efficiency of the pre-heat train is needed to raise the temperature of the crude oil before it goes into the furnace. This lowers the energy demand inside the furnace, which lowers the overall energy cost of the refinery. The crude pre-heat train and some utility heat exchangers are part of the heat exchanger network (HEN) of a crude distillation unit. A unique HEN is the pre-heat train, which is made up of many heat exchanges that are connected so that as much heat as possible can move between them [5]. Both the management of costs and the management of energy usage are significantly impacted by the analysis of heat exchanger networks (HENs). This field has advanced rapidly in recent years [6]. Assessment of energy losses and improvement potential falls into two types. Traditional energy analysis uses the first law of thermodynamics to assess energy waste relative to energy input without considering energy quality or process potential [7]. Another uses exergy analysis to assess the process's thermodynamic improvement potential by combining the first and second laws of thermodynamics [8, 9].

The usage of these approaches has resulted in significant contributions being made by a large number of scholars. For a considerable amount of time, the petrochemical and refining sectors have made substantial use of pinch analysis with the aim of lowering thermal energy. This has been done in order to achieve the goal of reducing thermal energy. The implementation of pinch analysis techniques has resulted in beneficial outcomes for a variety of other industries as well, including the pulp and paper industry, the chemical industry, and the food and beverage industry [6]. Through the enhancement of energy recovery between the hot streams (HS) and cold streams (CS) within the process, the major objective of heat integration is to reduce the amount of dependency on external energy sources, which are more often known as utilities. Because the ideal degree of heat recovery is reliant on the minimum temperature differential in the heat exchangers, which is symbolized by the symbol ΔT_{min}, the pinch approach is utilized for the purpose of determining the optimal degree of heat recovery. Using this method, one may determine the amount of heat that can be transferred between the hot and cold streams that are contained within the system, as well as the amount of cold and hot utilities that are required and the minimum energy requirements (MER).

The amount of energy that is accessible to be utilized is what is meant by the term "exergy." Exergy is a feature of thermodynamics that measures the greatest amount of valuable work that may be obtained from a system when it reaches a state of thermodynamic equilibrium with its surroundings [10]. The exergy analysis method has significant advantages over standard energy analysis since it can pinpoint the precise locations and types of irreversibilities that occur inside a specific system [11]. Exergy is always diminished when an irreversible event takes place. The dissipation of exergy drives the process of heat transfer. Therefore, it is more pragmatic to assess exergy efficiency using the principles of the second law of thermodynamics rather than relying on conventional energy efficiency [12].

1.2 Objectives

The objective of this thesis is to apply pinch, exergy and economic analysis to optimize the energy consumption of a distillation column in a petroleum refinery. The study aims to design an energy-efficient distillation column that meets the product specifications while minimizing energy consumption.

- To review the literature on energy efficiency in distillation columns and the application of pinch analysis in the petroleum refining industry.
- To simulate the distillation column process in HYSYS, and to validate the simulation against real-world data.
- To apply pinch analysis using Aspen Energy Analyzer to the distillation column process and to identify energy-saving opportunities.
- To apply exergy analysis to find exergy destruction in individual heat exchangers of the preheat train of the crude distillation unit.
- To find the exergy efficiency of an individual heat exchanger and overall heat exchanger network for the original crude distillation HYSYS model.

- To design an energy-efficient distillation column using the results of the pinch analysis and to compare the energy consumption of the optimized design with the original design.
- To evaluate the economic feasibility of the energy-efficient design, taking into account the capital and operating costs associated with the new design.
- To provide recommendations for future research on energy efficiency in distillation columns and the application of pinch analysis in the petroleum refining industry.

1.3 Organization of the Thesis

The thesis is organized into five chapters. Chapter 1 provides an introduction to the study, including the background, objectives, and scope of the study. Chapter 2 reviews the literature related to pinch, exergy analysis and its application in the energy optimization of distillation columns. Chapter 3 describes the methodology used for the study, including process description, process simulation, pinch analysis, exergy analysis formulations, energy-efficient design, simulation validation, and summary. Chapter 4 presents the results and discussions, including energy optimization, comparison of the base case and optimized designs, economic analysis, discussion of results, limitations, and summary. Chapter 5 provides a summary of the study, including contributions, future work, and conclusion. The references and appendices are provided at the end of the thesis.

CHAPTER 2: LITERATURE REVIEW

The crude distillation unit, often known as the CDU, is among the most important processing units in the refinery because of the significant amount of processing capacity it possesses. A crude oil heater or furnace, a network of heat exchangers, and distillation columns are the primary components that make up the unit. CDU is responsible for the separation of crude oil into its constituent parts, which include naphtha, kerosene oil, diesel, and other fractions with the necessary boiling ranges. As part of the process of fractionation, the stream of crude oil must be heated in the furnace to the right temperature in order to get the desired results. As a consequence of this, the conventional method of refining crude oil in CDUs necessitates the utilization of a significant quantity of heat energy and stripping steam. The operators of petroleum refineries are always working to improve the operational efficiency and energy utilization of their Crude Distillation Units (CDUs) in order to maximize their gross margins and reduce their carbon emissions [13, 14].

Increasing the temperature of crude oil before it enters the furnace is frequently required in order to reduce the amount of energy that is required inside the furnace. This is accomplished by the utilization of a series of heat exchangers, which are collectively referred to as the crude preheat train. Through the utilization of heat integration, the crude preheat train is produced. The process of preheating crude oil involves making use of the heat that is generated by hot products and pumping around the distillation column. The overall heat exchanger network of a crude distillation unit consists of a crude pre-heat train in addition to a few heat exchangers for utilities. Before the crude oil is heated up in the furnace, it is warmed, and the hot products of the distillation process are cooled down before they are discharged from the central distillation unit (CDU).

The analysis of heat exchanger networks (HENs) is an essential component in the process of optimizing the cost and expenditure of energy. Researchers and designers have used a number of different strategies in order to improve the integration of heat exchanger networks (HEN) and distillation, as well as to improve the design and retrofitting of HEN with the expectation of lowering the amount of energy that is consumed. This can be accomplished by either lowering the amount of energy that is required for the process or increasing the amount of energy that is recovered from the process [13]. HEN analysis, which identifies energy losses and improvement opportunities, falls into two kinds. Traditional energy analysis uses only the first law of thermodynamics. The other uses exergy analysis to assess the process's thermodynamic improvement potential by combining the first and second laws of thermodynamics [14].

In relation to the heat exchangers (HENs) of refineries, a number of research papers that are founded on the first law of thermodynamics have been documented. The majority of the research methodologies that have been used in the past for HEN analysis are based on pinch analysis methods. For instance, Mehdizadeh-Fard et al. [15] analyzed a real-life case study that involved a natural gas refinery that was quite complicated. For the purpose of improving heat recovery and optimizing the overall cold and hot utilities for the main processing units, the pinch analysis methodology was utilized. For this investigation, the methodology that was utilized consisted of the utilization of a comprehensive approach known as super-targeting, which centered on the overall heat transfer area and cost. Two primary approaches were developed: the "overall pinch method" for handling both the cold and hot process streams collectively and the "zonal targeting method" for targeting five separate locations inside the refinery, each of which contains some processing units. Both of these approaches were developed in order to address the cold and hot process streams. Following that, a comparative analysis was carried out in order to evaluate the results, and determinations were made regarding the possibilities for improvement in both instances. The results indicated that the application of the "zonal targeting method" in the modified HEN led to a considerable improvement in the overall energy utilization as well as the heat recoveries.

The crude distillation unit's energy consumption has also been reduced by using the pinch analysis approach. Ajao et al. [16] performed the energy integration of the crude pre-heat train of CDU I of the Kaduna Refinery and Petrochemicals business by using pinch analysis. A total cost index of 0.208 cost/s was needed. It was discovered that the temperature at the pinch point was 220 °C to reach the ideal minimum approach temperature of 15 °C. The utility targets were determined to be 1.112 x 108 kJ/hr for hot utilities and 1.018 x 108 kJ/hr for cold utilities, respectively, for the minimal approach temperature. A total of 38 heat exchangers were required to get the maximum amount of energy recovery.

Similarly, Al-Mutairi et al. [17] conducted a study on the heat integration and retrofitting of a Heat Exchanger Network (HEN) in a Crude Distillation Unit (CDU). The existing design was analyzed using pinch analysis, which identified opportunities for incorporating a retrofitting project into the design. The pinch analysis technique was employed to select the heat exchanger network (HEN) that maximized energy recovery inside the process. A variety of design options were explored and revised, finally leading to the selection of the most optimal design. The economic evaluation of the selected design indicated a decrease in the use of hot utilities and cold utilities by 8.4% and 10.9%, respectively, compared to the current design. Moreover, a yearly reduction in energy expenses amounting to $259,860 was accomplished. The project's payback period, which involved adapting the existing HEN to accommodate the newly redesigned design, was calculated to be eleven months.

In another study, Bulasara et al. [18] conducted a comprehensive analysis of the Heat Exchanger Network (HEN) in the Crude Distillation Unit (CDU) of a real refinery. This study examined the impact of including the readily accessible uncontrolled high-temperature streams from the Delayed Coking Unit (DCU). The study examined two specific scenarios for renovating using the pinch design method: (a) installing new heat exchangers for the entire network and (b) repurposing the existing heat exchangers. The objective of this study was to assess the feasibility of integrating the available high-temperature streams from the DCU section for heat use. In addition, the profitability of these streams was assessed when they were thermally integrated with the CDU in conjunction with the revamp study. A fresh graphical approach was created by Mamdouh A. Gadalla [19] for the purpose of representing an existing heat exchanger network or crude preheat train. For the purpose of illustrating the specifics of exchangers, the temperatures of cold streams were plotted against the temperatures of hot streams. In accordance with the principles of pinch analysis, the newly generated graphs made it easier to conduct a study and evaluation of the performance of the existing HENs. Due to the fact that the study identified energy inefficiencies in the current HEN, the next step was to conduct a statistical analysis of these inefficiencies in order to determine whether or not it is possible to recover energy. The graphical representation method that Gadalla proposed was utilized by Alhajri et al. [20] in order to adapt an existing HEN with the intention of optimizing the operations of a crude oil distillation process. For the purpose of conducting an investigation into a case study concerning a petroleum refinery located in Kuwait, the graphical approach was utilized. One of the key objectives of this inquiry was to carry out an energy analysis and make recommendations for retrofitting methods for the existing HEN. Based on the findings of the study, the minimal levels of utility use were identified, and potential energy and cost reductions were offered. When compared to the operational procedures that are now in place, the proposed method indicated that it was possible to achieve a potential energy savings of approximately 27%.

Researchers are showing an interest in exergy analysis, which is founded on the second law of thermodynamics. This is due to the fact that it has the capability of measuring the irreversibility of a process and evaluating its potential for development. Additionally, pinch analysis has been utilized in conjunction with exergy analysis in order to examine the HENs. In order to find the optimal minimum approach temperature ($\Delta Tmin$), Zun-long et al. [21] conducted an exergy-economic analysis of HEN systems for two case studies. The balanced composite curves were used in conjunction with the pinch analysis to determine the amount of energy that was consumed. As an alternative to the cost of utilities, the expense of exergy consumption was taken into consideration as the operating cost. This decision is made in conjunction with the capital cost in order to ascertain the ideal $\Delta Tmin$ for the synthesis of HENs.

Multiple investigations have been conducted on the exergy analysis of Heat Exchanger Networks in refineries. As an illustration, Mehdizadeh-Fard et al. [12] conducted an exergy analysis of the Heat Exchanger Network (HEN) at a complex natural gas refinery situated in the South Pars gas field. The advanced exergy analysis approach was applied to the serving heat exchanger network (HEN) to evaluate the possibilities for improving energy efficiency in the system. This study examined the concepts of the second law of thermodynamics, along with techno-economic limitations, in order to minimize the loss of available energy and maximize energy efficiency. Calculations were performed to determine both preventable and inevitable irreversibilities for each heat exchanger inside the plant's network. The analysis revealed that the overall efficiency of the Heat Exchanger Network (HEN) in the facility was 62.8%. However, it has the potential to be enhanced up to 84.2%, indicating a significant opportunity for development.

Exergy analysis has been used to evaluate the crude distillation unit and its heat exchanger network (HEN). Benali et al. [22] described a specific method of improving energy integration by modifying the flowsheet of a CDU. The application of exergy analysis was used to visually display and enhance comprehension of the distribution of energy degradations inside the distillation column. The objective was to identify a suitable solution for reducing these degradations. The observed distribution exhibited the preventable occurrence of lighter species across the whole distillation column. A suggestion was made to incorporate a pre-flash step into the pre-heating phase of the distillation process, in addition to effectively introducing the ascending vapors into the column. This might potentially lead to substantial energy conservation. These gains were attained by decreasing the furnace workload, implementing preparatory fractionation, and so reducing the irreversibilities of the column.

Another study by Izyan et al. [23] looked at how to lower a CDU's fuel usage using exergy analysis. It was found that the furnace was destroying the most exergy. Thus, two ways to reduce this loss were suggested: one was to raise the furnace's intake temperature, and the other was to reduce the amount of heat lost in the furnace's stack. Establishing suitable cleaning schedules for heat exchangers was also proposed.

Exergy analysis of a CDU's crude preheat train was also conducted by Fajardo et al. [2] through a case study investigation. In order to find the crucial regions and prospective areas for improvement, we used both standard and sophisticated exergy evaluations. According to the results, the HEN's total exergy destruction was more than 61.6 MW, with around 63% of that amount being listed as avoidable exergy destruction. Because they accounted for 39% of the total exergy degradation in the network, five heat exchangers were identified as crucial. In order to improve performance evaluation, we also looked at exergy efficiency to see how inevitable exergy destruction affected the rating of the exchanger's performance.

This study focuses on optimizing the crude preheat train within a petroleum refinery's crude distillation unit (CDU) through the integration of Exergy Analysis, Pinch Analysis (PA), and Economic Analysis for the developed model in Aspen HYSYS. Exergy Analysis is employed to identify energy inefficiencies within the heat exchanger network that open an area of improvement for energy conservation purposes. Pinch analysis application using an Aspen energy analyzer will create a more optimized heat exchanger network for crude distillation units in petroleum refineries. Also, a detailed economic analysis is to be done for the already developed simulation model for the crude distillation unit, and ultimately comparative analysis based on economics will be done for both base and optimized case heat exchanger network.

CHAPTER 3: METHODOLOGY

3.1 Process description

A schematic representation in Figure 2 depicts a crude distillation portion of a petroleum refinery. The flowsheet is based on the Aspen HYSYS model of a simplified crude distillation column that includes a Heat Exchanger Network that includes a crude pre-heat train and a few heat exchangers for utilities. To separate the light naphtha and gases from the heavy components, the first phase of the crude pre-heat train involves heating the cold crude oil to 232 degrees Celsius before sending it to the pre-flash drum. This is done in order to remove the substances. Before being transferred to the crude furnace, the bottom of the pre-flash drum is heated to a temperature of 279 degrees Celsius by the second section of the preheat train. The crude oil that has been heated in the furnace is then sent to the crude distillation column, which is set at a temperature of 343 degrees Celsius. Once there, it is separated into several straight-run fractions, such as naphtha, kerosene, diesel, and gas oil, amongst others. For the purpose of modeling the entire Heat Exchanger Network, a sub-flowsheet denoted as "Preheat Train" has been utilized. As a result of the close thermal coupling that exists between the HEN model and the crude distillation column, the simulation incorporates two dummy heaters that are referred to as "HEN-1" and "HEN-2." These heaters are specifically designed to replicate the first and second parts of the preheat train, respectively. Additionally, the overall topology of the whole Heat Exchanger Network is depicted in Figure 3, which also displays the exchanger matches that exist between the cold and hot process streams combined. According to the diagram, the network is comprised of twenty-one shell and tube heat exchangers that are interconnected with one another in order to lessen the amount of heat input that is necessary for the furnace.

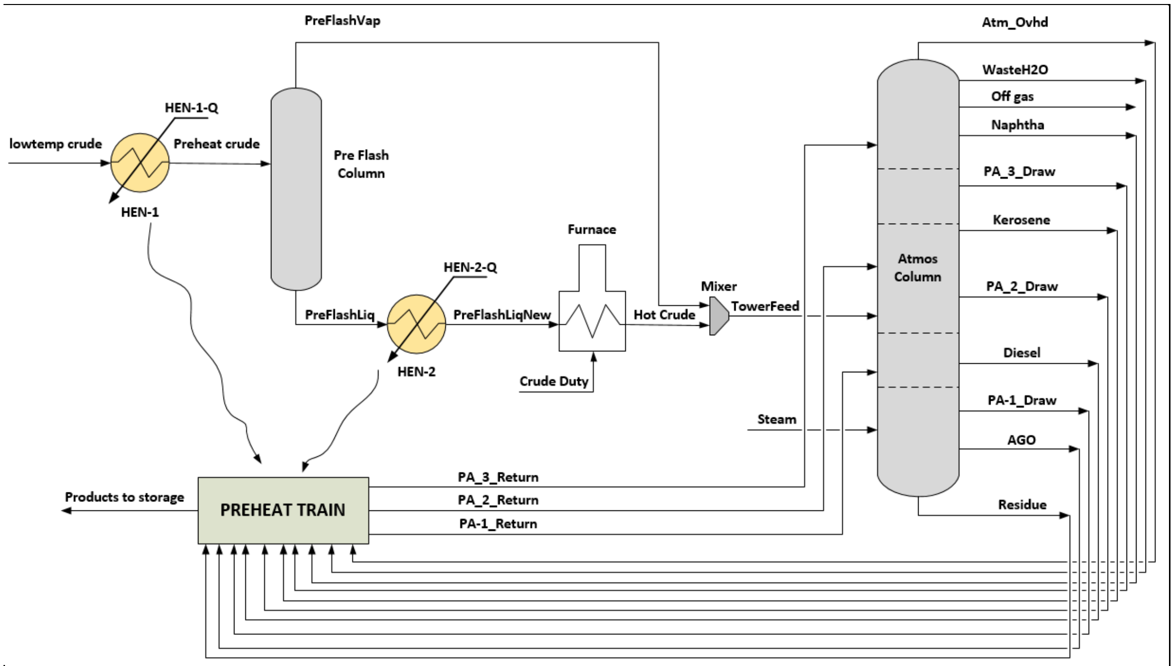


Figure 2: Flowsheet of a crude distillation unit

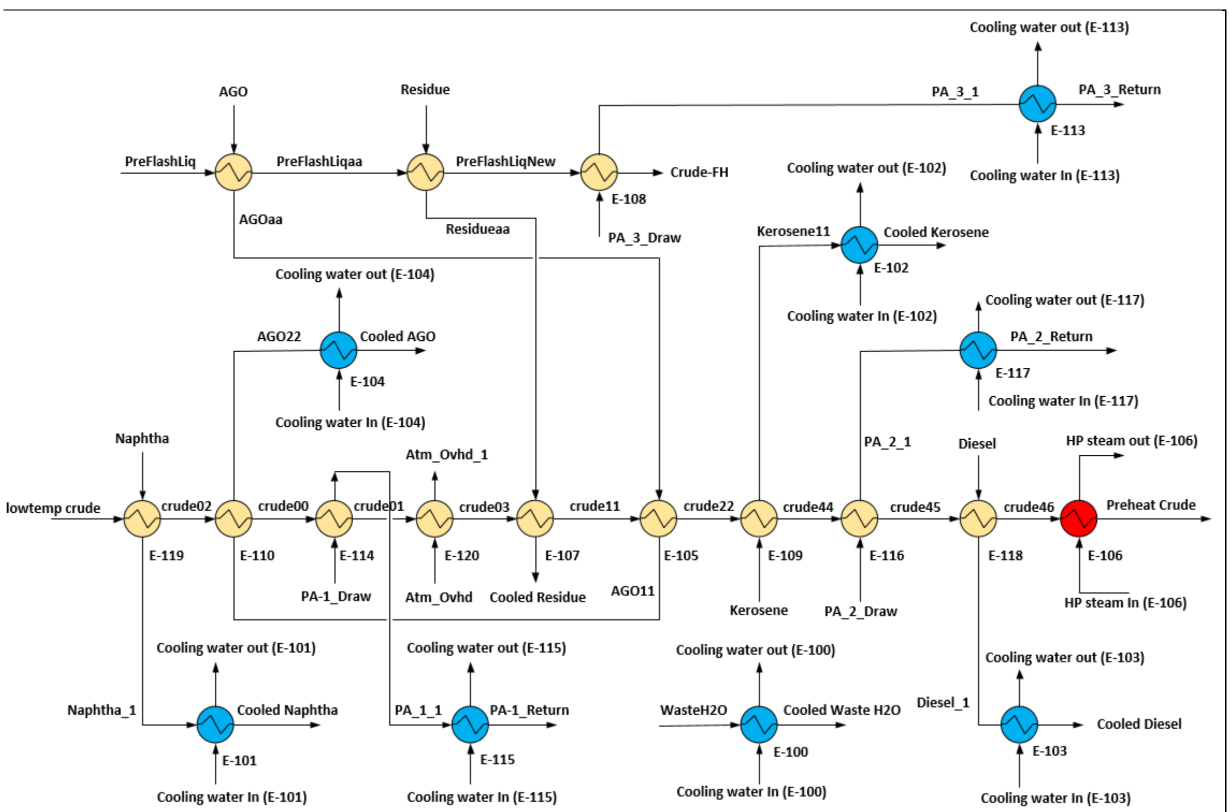


Figure 3: HEN of crude distillation unit

3.2 Methodology

This research is carried out in three phases, as shown in Figure 4 such as phase I, in which the crude distillation model is simulated in Aspen HYSYS along with exergy analysis of individual heat exchangers, pre-flash column, furnace, crude distillation unit, as well as overall exergy analysis of crude preheat train before crude distillation column. This analysis is done in HYSYS spreadsheets using appropriate data (mass flowrates and physical exergies) extraction of the inlet & outlet process (hot and cold) streams of every heat exchanger installed at crude preheat train and against every equipment. Also a exergy and number of trays trade off has been done for the selection of distillation column with minimum exergy destruction at optimal number of trays. In the second phase of the study, pinch analysis is done in the Aspen Energy Analyzer environment (AEA) environment for the efficient and optimized design of the heat exchanger network (HEN). Finally, in 3rd phase of this study, detailed economic analysis of the base and the optimized case is done using equipment direct costs from Aspen Process Economic Analyzer (APEA) and based on what production cost, profitability analysis, depreciation calculation and payback calculation is done for simulated crude distillation unit.

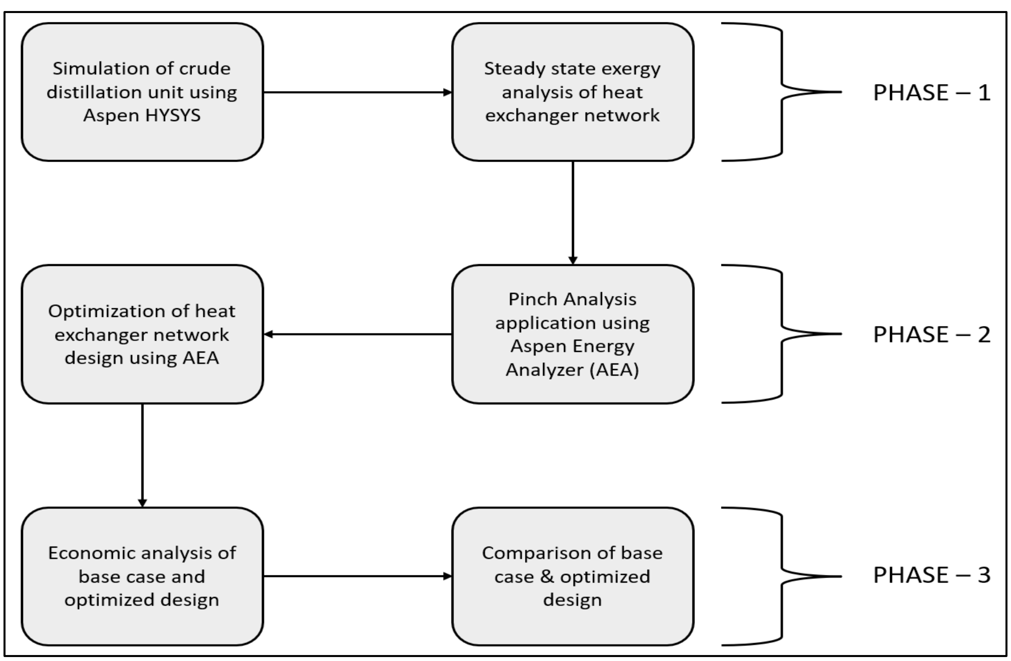


Figure 4: General phases of the study

3.3 Process Simulation

Aspen HYSYS environment is used to develop the crude distillation model where HEN-1 and HEN-2 represent the preheat train 1 (before pre-flash column) and preheat train 2 (before furnace heater of distillation column) as shown in figure 5. Both preheat trains are also developed as sub-flowsheets in the HYSYS environment, as shown in Figure 6. Figure 7 depicts the overall detailed picture of the distillation column along with side strippers of middle distillates such as diesel, kerosene and AGO.

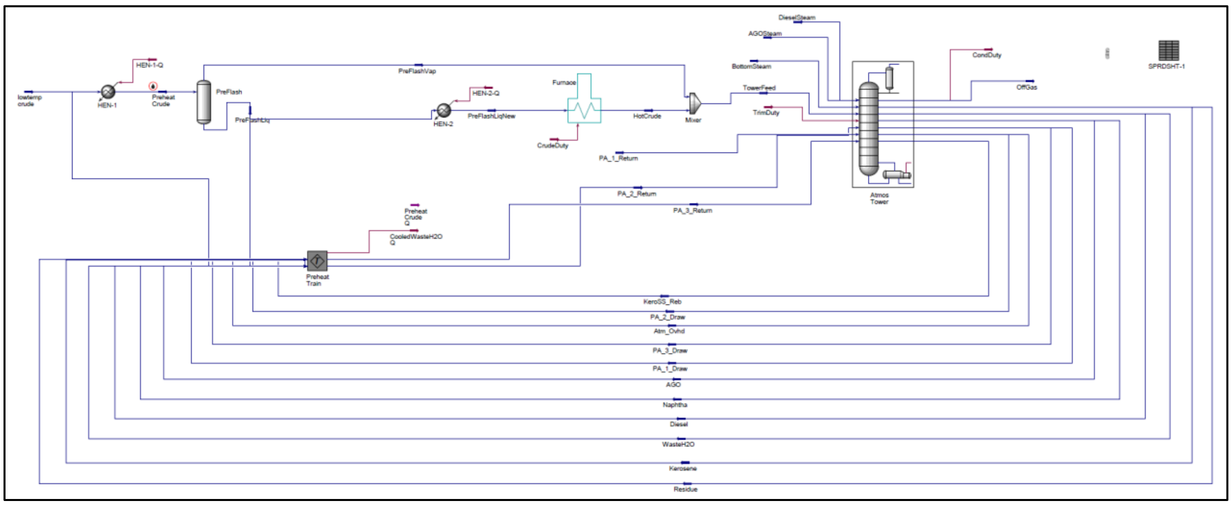


Figure 5: Flowsheet of crude distillation unit in HYSYS

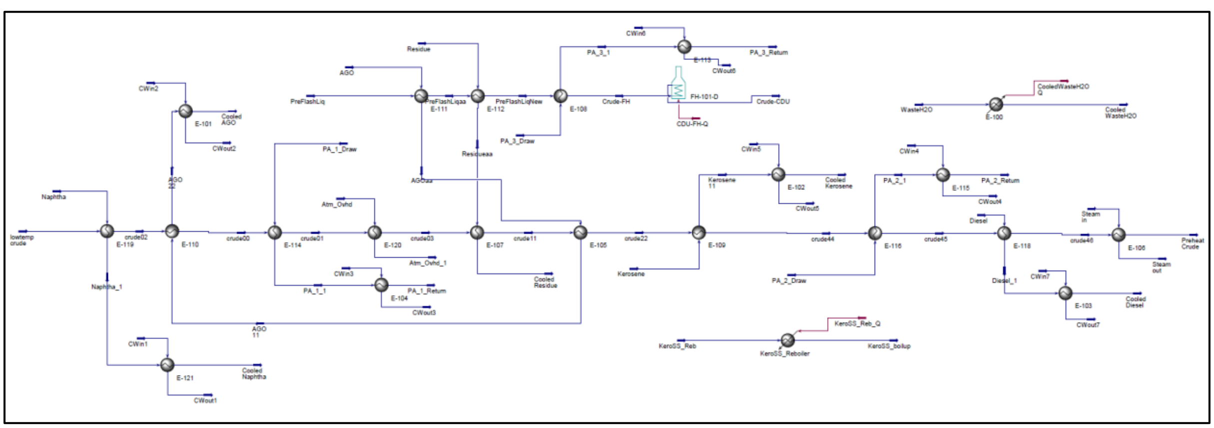


Figure 6: Crude preheat train (1 & 2) of crude distillation unit

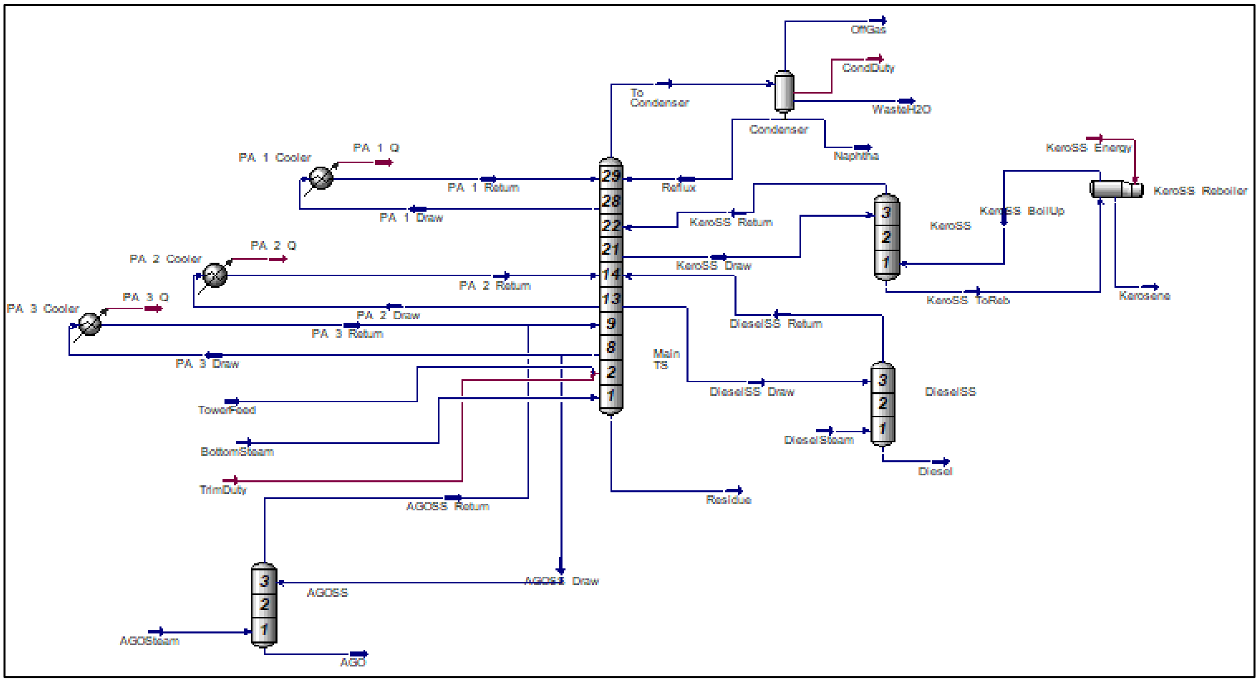


Figure 7: Overall HYSYS model of the distillation column and side strippers

3.3.1 Simulation Validation

Data from Crude distillation unit products and the HYSYS module were compared in order to validate the simulation. The comparative analysis findings demonstrated that the products' liquid volume flows differ significantly. Comparing the two indicated that the adaption measurement and retrofit could both be safely completed using the HYSYS program in the current study, as shown in Table 1.

Table 1: Comparison between HYSYS results and actual crude distillation data

Product	**Literature (m^3/hr)**	**HYSYS Results (m^3/hr)**	**Error %**
Off gas	21.8625	21.87	0.000343
Naphtha	114.6125	114.6	0.000110
Kerosene	85.4625	85.46	0.000029

Diesel	113.95	113.90	0.000440
AGO	32.4625	32.46	0.000077
Residue	294.15	296.00	0.006250

3.4 Exergy Analysis Formulations

Exergy can be well-defined as “the system's useful energy”. It is the summation of physical, chemical, kinetic, and potential exergies.

$$E = E_{PH} + E_{CH} + E_{KN} + E_{PT} \tag{1}$$

As there are no chemical reactions involved, any change in the exergy of the HEN is due to physical changes. Furthermore, the changes in potential exergy and kinetic exergy can also be ignored.

3.4.1. Exergy Analysis of Heat Exchangers

The analysis of thermodynamics second law can be used to examine the distribution of exergy load inside a heat exchanger or heat exchanger network containing cold and hot process streams, to study the process of heat exchange. The values of exergy for all the input and output streams must be computed to perform an exergy balance on a heat exchanger. The following steady-state equations can be employed to perform the exergy balance of any heat exchanger.

$$I_{ex} = E_d = \left(E_{H,in} - E_{H,out}\right) + \left(E_{C,in} - E_{C,out}\right) = T_o S_{gen} \tag{2}$$

Exergy efficiency is a quantitative parameter employed to assess how effectively the exchange of exergy is occurring between cold and hot process streams of a heat-transfer process. The exergy efficacy of a heat exchanger can be stated as "the ratio of exergy produced to the exergy consumed”. The exergy content of one process fluid rises while the exergy content of the other decreases in the case of heat exchangers. Thus, the hot process stream or cold process stream will release or consume the exergy accordingly.

$$\eta_{ex} = \frac{E_p}{E_c} = \frac{(E_{C,out} - E_{C,in})}{(E_{H,in} - E_{H,out})} \quad For\ T_H \& T_C > T_o \tag{3}$$

$$\eta_{ex} = \frac{E_p}{E_c} = \frac{(E_{H,out} - E_{H,in})}{(E_{C,in} - E_{C,out})} \quad For\ T_H \& T_C < T_o \tag{4}$$

The exergy content of the cold process stream reduces rather than improves in case of heat transfer occurring across the ambient temperature. Because the stream creates no exergy or very minimum exergy, the efficiency will also be zero or extremely low.

3.4.2. Exergy Analysis of Heat Exchanger Network

In the case of a heat exchanger network, the sum of exergy destructions for all those specific heat exchangers operating in the network can be used to compute the overall irreversibility caused due to the total heat transferred:

$$I_{HEN} = \Sigma E_d = \Sigma E_{in} - \Sigma E_{out} = \Sigma\left(E_{H,in} - E_{H,out}\right) + \Sigma\left(E_{C,in} - E_{C,out}\right) \tag{5}$$

Also, the overall exergy efficiency of the entire heat exchanger network can be calculated as:

$$\eta_{HEN} = \frac{\Sigma E_p}{\Sigma E_c} \tag{6}$$

The following suppositions were made during the process of performing exergy analysis.

- Heat exchanger units were modeled and evaluated as a steady-state flow system.
- The values of potential and kinetic exergies were neglected.
- The reference conditions for exergy calculations were established at a temperature of 25°C and a pressure of 101.325 kPa.

The values of physical exergies of the process streams were computed using the property set of Aspen HYSYS. Then using the values of physical exergy, irreversibilities and exergy efficiencies of individual heat exchangers were determined by using equations (2) and (3).

Furthermore, equations (5) and (6) were used for the calculation of the irreversibilities and exergy efficiency of the whole heat exchanger network. Figure 9 shows the general flow scheme of the system's exergy showing that exergy is being wasted due to irreversibility inside the system.

3.4.3. Exergy Analysis of Preflash column

For the preflash unit, the exergy analysis focuses on identifying the exergy destruction due to phase separation and temperature changes in the unit. The exergy balance for the preflash unit can be written as:

$$I_{preflash} = Ex_{in} - Ex_{out} = E_{destruction} \qquad (7)$$

Where:

- Ex_{in} is the total exergy input to the preflash unit.
- Ex_{out} is the exergy output in both vapor and liquid streams.
- $Ex_{destruction}$ is the exergy destroyed due to irreversibilities in the preflash process.

The physical exergy of the streams entering and leaving the preflash unit is given by:

$$Ex_{physical} = (h - ho) - To(S - So) \qquad (8)$$

Where:

- h and S are the specific enthalpy and entropy of the stream.
- ho and So are the specific enthalpy and entropy of the stream at the reference state.
- To is the ambient temperature.

The exergy efficiency of the preflash unit is given by:

$$\eta_{ex,preflash} = \frac{E_{xout}}{E_{xin}} \qquad (9)$$

3.4.4. *Exergy Analysis of Furnace*

In the heater, energy is supplied to the process, but exergy is lost due to irreversibilities, including heat losses and entropy generation. The exergy analysis for the heater can be described by the exergy destruction due to heat transfer at a finite temperature difference.

The exergy destruction in the heater is:

$$Ex_{destruction,furnace} = Ex_{in} - Ex_{out} - \frac{Q}{T}(To - Theater) \qquad (10)$$

Where:

- Ex_{in} is the total exergy input to the heater.
- Ex_{out} is the exergy leaving heater.
- T_{heater} is the temperature at which the heat is supplied.

The exergy efficiency of the heater is:

$$\eta_{ex,heater} = \frac{E_{xout}}{E_{xin}} \qquad (11)$$

3.4.5. *Exergy Analysis of crude distillation column*

For the crude distillation column, exergy analysis focuses on identifying exergy destruction across the column due to mass transfer, heat transfer, and pressure changes.

The overall exergy destruction in the distillation column is:

$$Ex_{destruction,column} = \sum(Ex_{in,streams} - Ex_{out,streams}) \qquad (12)$$

Where:

- $Ex_{in,streams}$ is the exergy of all streams entering the distillation column (feed, reboiler heat, etc.).
- $Ex_{out,streams}$ is the exergy of all streams leaving the distillation column (products, condenser heat, etc.).

The exergy efficiency of the distillation column is:

$$\eta_{ex,column} = \frac{E_{xout,products}}{E_{xin,feed}} \tag{13}$$

This measures how effectively the distillation column converts the exergy of the feed into useful product streams.

Equations 7 to 13 are used for the calculation of exergy destruction and exergy efficiency across individual equipment.

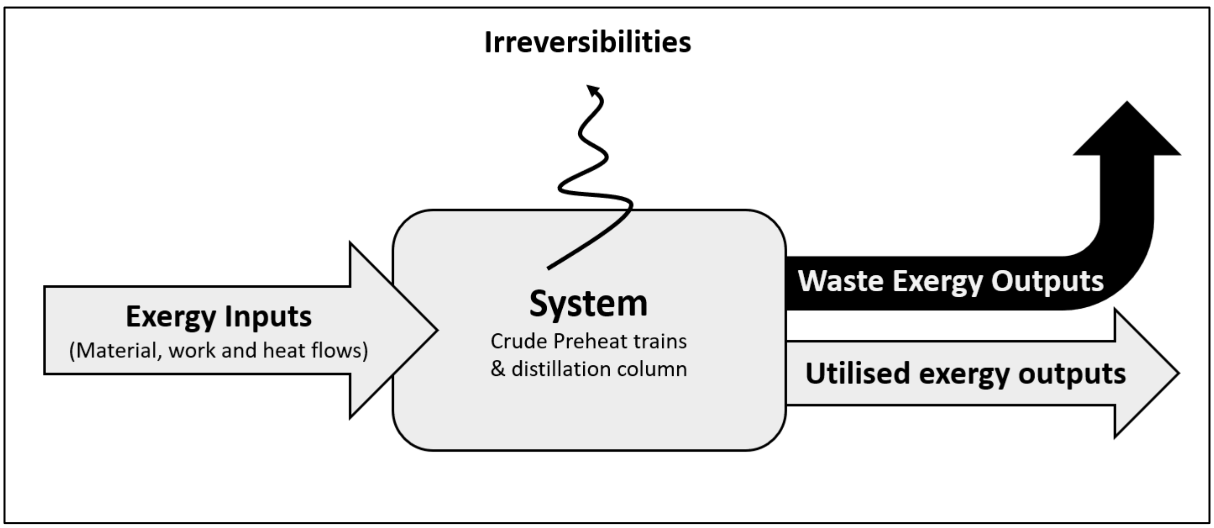


Figure 8: General exergy balance of the system

3.5 Pinch Analysis

3.5.1. Pinch analysis sequence

The steps that are followed in pinch analysis are outlined in Figure 9. A pinch technology study was conducted for the crude distillation process flow diagram. Following the identification of process streams, data is retrieved from the steady-state process flow diagram in the first step of the analysis, and the grid diagram is created. The refinery unit's design and operational data were derived from the process simulation. In order to perform a convergence test and validate the stream enthalpy data, an extensive modeling and simulation of the unit was conducted in the Aspen HYSYS environment. The required enthalpy data were extracted. This was carried out in order to assess the current network for possible cost and energy savings.

Moreover, the targets for the hot and cold utilities, as well as the pinch point, were determined by analyzing the thermodynamic profiles of the process streams utilizing the Problem Table methods, Composite Curves, and Grand Composite Curves. This profile also indicated the maximum energy recovery possible at the chosen ΔTmin. Aspen Energy Analyzer analyzed the pinch technology of this process.

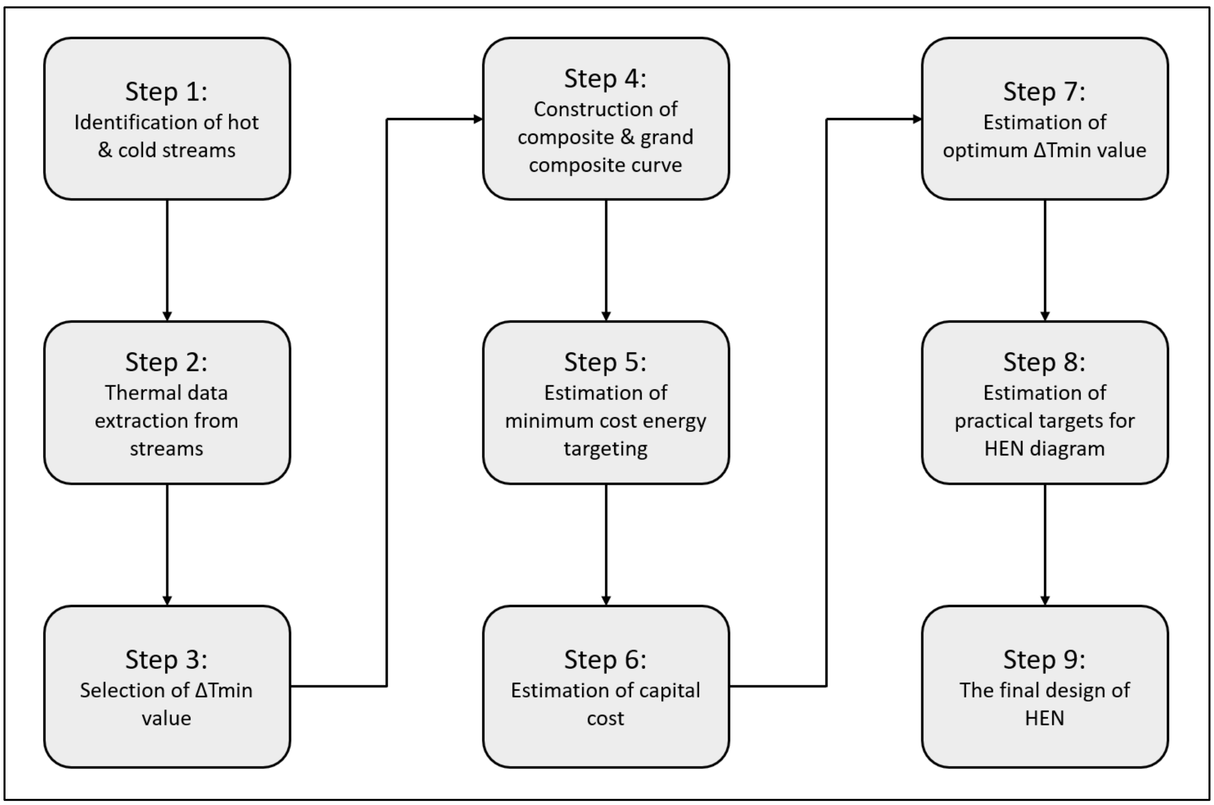


Figure 9: Pinch Analysis Steps

3.5.2. Process & utility streams data

The pinch occurs where the hot streams are at 231.8 °C and the cold at 221.8 °C**.** The pinch is depicted as a vertical continuous black line in Figure 10 below, which displays the grid structure for this process. The above and below pinch designs are the two separate portions that make up the crude refining heat exchanger network design.

Table 2: Process Hot and Cold Stream Data for Heat Network Design

S#	Stream Name	Inlet T (°C)	Outlet T (°C)	MCp (kJ/C-h)	Enthalpy (kJ/hr)
1	KeroSS_Reb_To_KeroSS_boiler	226.2	240.2	---	22033256
2	Atm_Ovhd_To_Atm_Ovhd1	146.5	110	---	28668504
3	WasteH2O_To_CooledWasteH2O	73.2	40	23819	790779
4	Diesel_To_Cooled Diesel TPL1	248.5	50	---	45828649
5	Naphtha_To_Cooled Naphtha@TPL1	73.2	30	212173	9165870
6	PA_3_Draw_To_PA_3_Return	319.6	244.1	488891	36911249
7	PA_1_Draw_To_PA_1__Return	166.9	71.7	---	58033087
8	AGO_To_Cooled AGO@TPL1	298	70	---	16717222
9	PA_2_Draw_To_PA_2_Return	263	180.5	444815	36697261
10	Lowtemp crude_To_Preheat crude	30	235	---	281285150
11	Preflashliq_To_Crude-CDU@TPL1	232.2	343	---	193816834
12	Kerosene_To_Cooled Kerosene@TPL1	231.8	40	---	31987137
13	Residue_To_Cooled Residue@TPL1	347.3	120	---	169952744

The following grid diagram has been created with an Aspen Energy Analyzer using ΔTmin = 10°C, showing process streams above and below the pinch point.

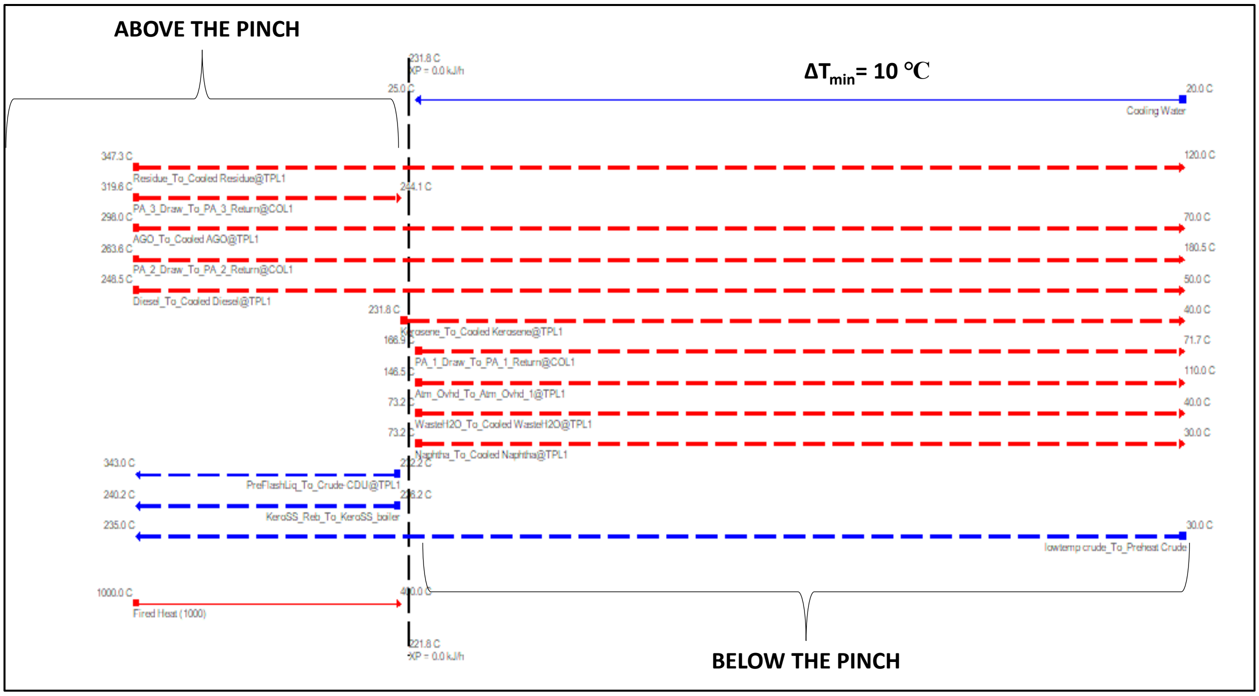


Figure 10: Grid diagram displaying the pinch segments above and below the pinch point

Name		Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
KeroSS_Reb_To_KeroSS_b		226.2	240.2	---	2.203e+007		----	9.776e+004	----	Global
Atm_Ovhd_To_Atm_Ovhd1		146.5	110.0	---	2.867e+007		----	1.315e+005	----	Global
WasteH2O_To_CooledWasl		73.2	40.0	2.382e+004	7.908e+005		18627.31	5686	4.189	Global
Diesel_To_Cooled Diesel TF		248.5	50.0	---	4.583e+007		----	8.614e+004	----	Global
Naphtha_To_Cooled Naphth		73.2	30.0	2.122e+005	9.166e+006		4496.71	9.181e+004	2.311	Global
PA_3_Draw_To_PA_3_Retu		319.6	244.1	4.889e+005	3.691e+007		4909.63	1.551e+005	3.152	Global
PA_1_Draw_To_PA_1__Re		166.9	71.7	---	5.803e+007		----	2.391e+005	----	Global
AGO_To_Cooled AGO@TPl		298.0	70.0	---	1.672e+007		----	2.603e+004	----	Global
PA_2_Draw_To_PA_2_Retu		263.0	180.5	4.448e+005	3.670e+007		5037.52	1.512e+005	2.942	Global
lowtemp crude_To_Preheat		30.0	235.0	---	2.813e+008		----	5.190e+005	----	Global
Preflashliq_To_Crude-CDU@		232.2	343.0	---	1.938e+008		----	4.964e+005	----	Global
Kerosene_To_Cooled Kerosi		231.8	40.0	---	3.199e+007		----	6.399e+004	----	Global
Residue_To_Cooled Residu		347.3	120.0	---	1.700e+008		----	2.512e+005	----	Global

Figure 11: Process streams data in the AEA environment

Cooling water and fired heat of 1000 °C were selected as cooling and heating utilities for completing the heat exchanger network.

Name		Inlet T [C]	Outlet T [C]	Cost Index [Cost/kJ]	Segm.	HTC [kJ/h-m2-C]	Target Load [kJ/h]	Effective Cp [kJ/kg-C]	Target FlowRate [kg/h]	DT Cont. [C]
Cooling Water		20.00	25.00	2.125e-007		12504.76	2.251e+007	4.183	1076322.63	5.00
Fired Heat (1000)		1000	400.0	4.249e-006		399.60	8.489e+007	1.000	141490.04	25.00

Figure 12: Utility streams data in the AEA environment

3.5.3. *Targets*

After putting process and utility stream data, the software produces target values for the heat exchanger network, as shown in figure 13 below.

Heating [kJ/h]	8.489e+007
Cooling [kJ/h]	2.251e+007
Number of Units	14
Total Area [m2]	1.691e+004
Op. Cost Index [Cost/s]	0.1015
Capital Cost Index [Cost]	4.601e+006
Total Cost Index [Cost/s]	0.1400

Figure 13: Target values of heat exchanger network

3.5.4. *Energy-Efficient Design of the Distillation Column*

Figure 14 shows the maximum energy recovery (MER) design for the process streams to achieve their desired temperatures while exchanging heat between cold and hot streams. The rest of the rise and fall for cold and hot streams are done using hot and cold utilities. This design consists of 20 heat exchangers to manage heat transfer properly and is considered a base case heat exchanger network for crude distillation units.

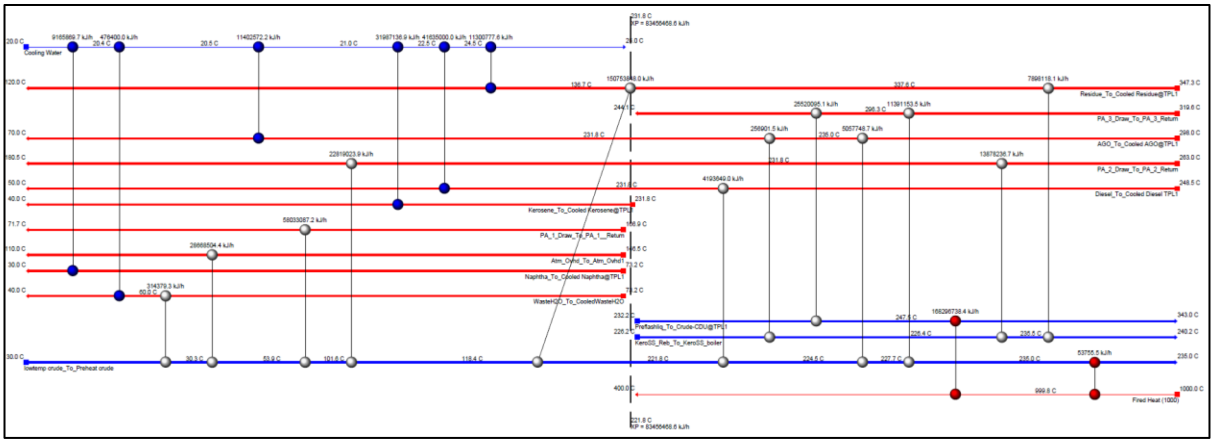


Figure 14: Base case heat exchanger network.

3.5.5. *Energy Optimization of the Distillation Column*

The basic purpose of pinch analysis application to the heat exchanger network is to optimize the energy transfer between process streams with less capital and operating cost. This purpose is fulfilled using an optimization tool in Aspen Energy Analyzer to develop a more optimized heat exchanger network using only 18 heat exchangers while keeping heating and cooling targets the same as per design. Figure 15 shows the optimized heat exchanger network for the pre-heat train.

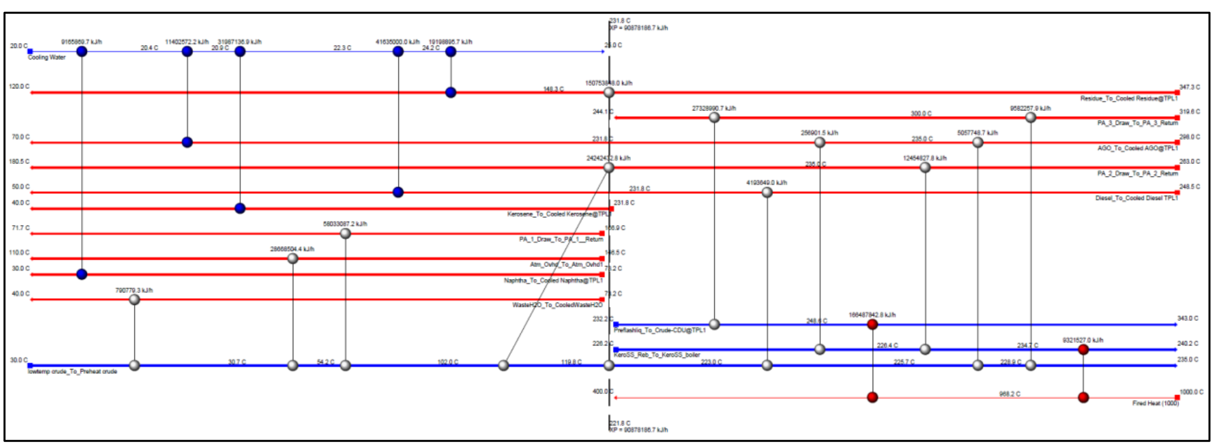


Figure 15: Optimized heat exchanger network (HEN)

CHAPTER 4: RESULTS AND DISCUSSION

Section 4.1 describes the exergy analysis results of the entire Heat Exchanger Network. In section 4.2, results (composite curves, grand composite curves, HEN design for base and optimized case, utility data for base and optimized case) of pinch analysis are discussed. Also, a comparative analysis of base and optimized design in terms of utility requirements of heat exchangers is discussed. Finally, in section 4.3, detailed results of economic analysis of the optimized design are discussed, followed by results discussion and study limitations.

4.1 Exergy Analysis

Table 3 presents the exergy calculations and other operating conditions of the process streams. Physical exergy values of the process streams were computed using the original simulation model developed in the Aspen HYSYS environment. The estimation of the chemical exergies of the process streams was not performed. The heat exchanger network of crude distillation units included twenty-one shell and tube heat exchangers. From the data of process streams, the exergy destructions and exergy efficiencies were determined for all the individual heat exchangers of the heat exchanger network, as shown in Table 4 and Table 5. Furthermore, the complete exergy destruction and the overall exergy efficiency for the whole HEN of the crude distillation unit were calculated.

Table 3: Operating Conditions and exergy data of process streams

Streams	**Mass flow rate (ton/hr)**	**Temperature (C)**	**Pressure (kPa)**	**Physical Exergy (kW)**
AGO	26.04	298.7	218.6	1641
AGO 11	26.04	211	218.6	787
AGO 22	26.04	120	218.6	210
AGOaa	26.04	262.2	218.6	1253
Atm_Ovhd	128.8	147.3	197.9	2928

Atm_Ovhd_1	128.8	110	197.9	854
Cooled AGO	26.04	70	218.6	48
Cooled Diesel	86.2	50	213.6	51
Cooled Kerosene	64.01	40	205.8	16
Cooled Naphtha	91.84	38	135.8	14
Cooled Residue	251.1	156.5	225.5	3822
Cooled Waste H2O	5.694	40	135.8	1.50
cooling water in (E-100)	39.04	35	101.3	3.1
cooling water in (E-101)	168.7	35	101.3	13.24
cooling water in (E-102)	1414	35	101.3	111
cooling water in (E-103)	1841	35	101.3	144
cooling water in (E-104)	154.2	35	101.3	12.1
cooling water in (E-113)	1041	35	101.3	82
cooling water in (E-115)	884.7	35	101.3	69.4
cooling water in (E-117)	411.3	35	101.3	32.3
cooling water out (E-100)	39.04	40.0	101.3	9.85
cooling water out (E-101)	168.7	38.0	101.3	29.2
cooling water out (E-102)	1414	39.8	101.3	350.3

cooling water out (E-103)	1841	39.9	101.3	457.6
cooling water out (E-104)	154.2	39.9	101.3	38.3
cooling water out (E-113)	1041	39.9	101.3	263
cooling water out (E-115)	884.7	40.0	101.3	223.4
cooling water out (E-117)	411.3	40.0	101.3	104.0
crude00	519	40.9	517.1	189.0
crude01	519	75	517.1	1279.8
crude02	519	35	517.1	113.8
crude03	519	98.2	517.1	2644.1
crude11	519	163	517.1	8928.6
crude22	519	165.8	517.1	9285.9
crude44	519	167.7	517.1	9522.4
crude45	519	187.3	517.1	12185.2
crude46	519	192.5	517.1	12928.8
Crude-FH	496.9	278.5	517.1	28173.1
Diesel	86.2	249.3	213.6	3685.4
Diesel_1	86.2	220	213.6	2823.5
HP steam in (E-106)	28.41	500	3000	10247.3
HP steam out (E-106)	28.41	234.5	3000	2406.6
Kerosene	64.01	233.2	205.8	2360.9

Kerosene 11	64.01	219	205.8	2065.3
low-temp crude	519	30	517.1	76.7
Naphtha	91.84	74.26	135.8	196.9
Naphtha_1	91.84	48.5	135.8	45.4
PA_1_1	239.2	103.1	198.9	1334.4
PA_1_Draw	239.2	167.8	198.9	4297.0
PA_1_Return	239.2	70.41	198.9	453.2
PA_2_1	151.3	201	213.6	4098.8
PA_2_Draw	151.3	264.2	213.6	7326.2
PA_2_Return	151.3	180.7	213.6	3243.5
PA_3_1	155.1	290	218.6	9209.6
PA_3_Draw	155.1	319.9	218.6	11246.0
PA_3_Return	155.1	245.2	218.6	6499.4
PreFlashLiq	496.9	232.2	517.1	18518.6
PreFlashLiqaa	496.9	234	517.1	18865.8
PreFlashLiqNew	496.9	270	517.1	26252.0
Preheat Crude	519	232.2	517.1	19844.8
Residue	251.1	347.4	225.5	21500.2
Residueaa	251.1	273.5	225.5	13170.2
WasteH2O	5.694	74.24	135.8	22

4.1.1. Exergy Destruction

Process irreversibility determines the measure of exergy destroyed in a unit operation. Or process. Irreversibility in any unit operation or process is caused by to:

1) Spontaneous chemical reaction

2) Depletion of work into heat by the solid or fluid friction.
3) Heat transfer at finite temperature differences
4) Unconstrained expansion or thermal equilibrium in a mixing [8, 24]

Table 4 represents the exergy destruction of individual heat exchangers and the overall exergy destruction of the entire Heat Exchanger Network. The value for overall exergy destruction of HEN was 17611.2 kW. Figure 16 represents the exergy destruction contributions of all heat exchangers to the overall exergy destruction of HEN. Heat exchangers predominantly contributing to the overall exergy destruction include E-107, E-113, E-103 and E-102, with the exergy destruction values of 3063.7 kW, 2529.1 kW, 2459.4 kW and 1810.1 kW respectively. Only these four out of twenty-one heat exchangers contribute 56 % to the overall exergy destruction of HEN. So, to enhance the exergy efficiency of HEN and achieve more significant energy savings, priority must be given to the optimization of these four heat exchangers. The primary purpose is to minimize the exergy destruction of these heat exchangers using energy optimization strategies.

Table 4: Exergy destruction of all the individual heat exchangers

Exchanger No.	**Inlet and outlet streams**				**Exergy Destruction (kW)**
	Hot in	**Hot out**	**Cold in**	**Cold out**	
E-100	WasteH2O	Cooled Waste H2O	cooling water in (E-100)	cooling water out (E-100)	13.7
E-101	Naphtha_1	Cooled Naphtha	cooling water in (E-101)	cooling water out (E-101)	15.4
E-102	Kerosene 11	Cooled Kerosene	cooling water in (E-102)	cooling water out (E-102)	1810.1
E-103	Diesel_1	Cooled Diesel	cooling water in (E-103)	cooling water out (E-103)	2459.4
E-104	AGO 22	Cooled AGO	cooling water in (E-104)	cooling water out (E-104)	135.0
E-105	AGOaa	AGO 11	crude11	crude22	107.9

E-106	HP steam in (E-106)	HP steam out (E-106)	crude46	Preheat Crude	924.7
E-107	residue	Cooled Residue	crude03	crude11	3063.7
E-108	PA_3_Draw	PA_3_1	PreFlashLiq New	Crude-FH	115.2
E-109	Kerosene	Kerosene 11	crude22	crude44	59.0
E-110	AGO 11	AGO 22	crude02	crude00	502.4
E-111	AGO	AGOaa	PreFlashLiq	PreFlashLiqaa	41.4
E-112	Residue	Residueaa	PreFlashLiqa a	PreFlashLiqN ew	943.8
E-113	PA_3_1	PA_3_Return	cooling water in (E-113)	cooling water out (E-113)	2529.1
E-114	PA_1_Draw	PA_1_1	crude00	crude01	1871.8
E-115	PA_1_1	PA_1_Return	cooling water in (E-115)	cooling water out (E-115)	727.2
E-116	PA_2_Draw	PA_2_1	crude44	crude45	564.5
E-117	PA_2_1	PA_2_Return	cooling water in (E-117)	cooling water out (E-117)	783.5
E-118	Diesel	Diesel_1	crude45	crude46	118.1
E-119	Naphtha	Naphtha_1	lowtemp crude	crude02	114.5
E-120	Atm_Ovhd	Atm_Ovhd_1	crude01	crude03	709.9

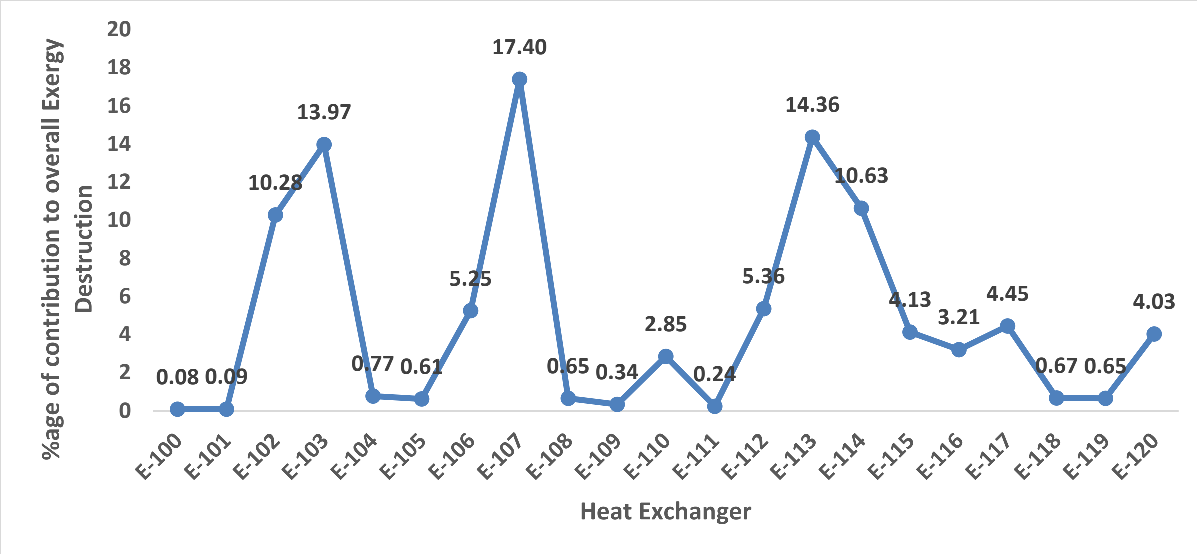


Figure 16: Contribution of heat exchangers to the overall exergy destruction of the entire HEN

4.1.2. Exergy Efficiency

The exergy efficiency quantifies the effectiveness of a system in relation to its performance. The overall exergy efficiency of the Heat Exchanger Network was 63.34%. Table 5 presents the exergy efficiencies of individual heat exchangers, calculated from equation (3). Among the heat exchangers, E-113 was least efficient, followed by E-117, E-103, E-102 and E-110 with exergy efficiency of 6.6%, 8.3%, 11.2%, 11.6% and 13.0% respectively. In comparison, E-108 has the highest exergy efficiency of 94.3%. Figure 17 shows the visual presentation of exergy efficiencies of individual heat exchangers belonging to the HEN. Effective heat exchanger design can limit the losses of heat and pressure [25, 26]. The exergy efficiency of heat exchangers can be improved by enhancing the thermal designs of heat exchangers, keeping in mind the techno-economic aspects.

Table 5: Exergy efficiencies of individual heat exchangers

Exchanger No.	**Inlet and outlet streams**				**Exergy Efficiency (%)**
	Hot in	**Hot out**	**Cold in**	**Cold out**	
E-100	WasteH2O	Cooled Waste H2O	cooling water in (E-100)	cooling water out (E-100)	33.1

E-101	Naphtha_1	Cooled Naphtha	cooling water in (E-101)	cooling water out (E-101)	50.8
E-102	Kerosene 11	Cooled Kerosene	cooling water in (E-102)	cooling water out (E-102)	11.6
E-103	Diesel_1	Cooled Diesel	cooling water in (E-103)	cooling water out (E-103)	11.2
E-104	AGO 22	Cooled AGO	cooling water in (E-104)	cooling water out (E-104)	16.2
E-105	AGOaa	AGO 11	crude11	crude22	76.7
E-106	HP steam in (E-106)	HP steam out (E-106)	crude46	Preheat Crude	88.2
E-107	Residueaa	Cooled Residue	crude03	crude11	67.2
E-108	PA_3_Draw	PA_3_1	PreFlashLiqNew	Crude-FH	94.3
E-109	Kerosene	Kerosene 11	crude22	crude44	80.0
E-110	AGO 11	AGO 22	crude02	crude00	13.0
E-111	AGO	AGOaa	PreFlashLiq	PreFlashLiqaa	89.3
E-112	Residue	Residueaa	PreFlashLiqaa	PreFlashLiqNew	88.6
E-113	PA_3_1	PA_3_Return	cooling water in (E-113)	cooling water out (E-113)	6.6
E-114	PA_1_Draw	PA_1_1	crude00	crude01	36.8
E-115	PA_1_1	PA_1_Return	cooling water in (E-115)	cooling water out (E-115)	17.4
E-116	PA_2_Draw	PA_2_1	crude44	crude45	82.5
E-117	PA_2_1	PA_2_Return	cooling water in (E-117)	cooling water out (E-117)	8.38

E-118	Diesel	Diesel_1	crude45	crude46	86.2
E-119	Naphtha	Naphtha_1	lowtemp crude	crude02	24.4
E-120	Atm_Ovhd	Atm_Ovhd_1	crude01	crude03	65.7

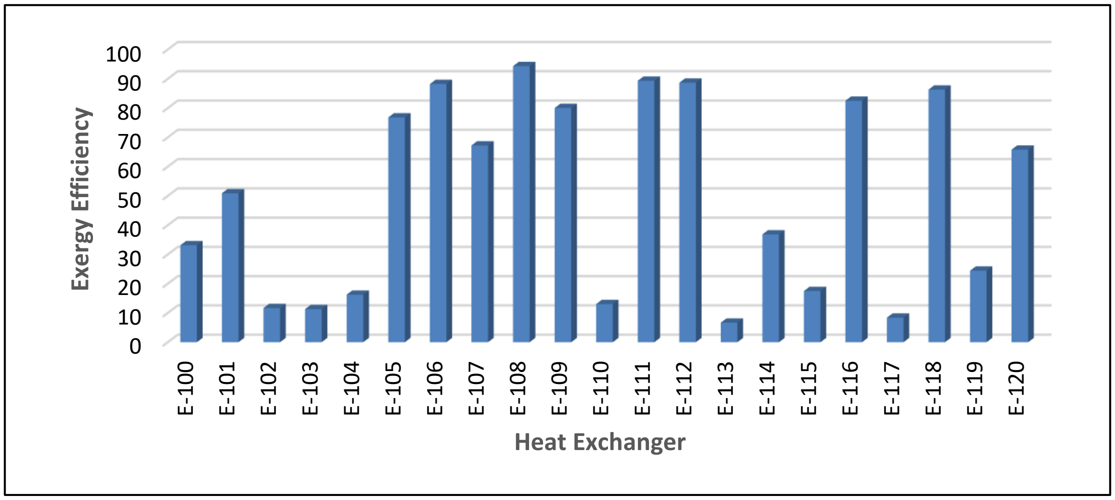


Figure 17: Exergy efficiencies of all the heat exchangers of HEN

4.1.3. Exergy Destruction and Efficiency of Other Process Units

The following table 6 shows the exergy efficiency and exergy destruction of each unit operations.

Table 6: Exergy efficiency and exergy destruction of all process units

Units	**Exergy Efficiency (%)**	**Exergy Destruction (kW)**	**% Contribution to total exergy destruction**

Preheat Train	63.34	17,611.2	3.10
Furnace	76.1	39,260.06	6.91
Preflash Drum	74.1	195,863.10	34.50
Crude Distillation Tower	53.1	314,670.11	55.43
Mixer	99.8	298.23	0.06
Total		567,702.69	100

It could be observed from the results presented in the above table that the atmospheric distillation tower has the lowest exergy efficiency of 53.1% followed by the pre-flash drum (74.1%) and furnace (76.1%). The heat exchangers network, however, also have lower exergy efficiency in the range of 63.34%, while the mixer has the highest exergy efficiency of 99.9%. The column has the lowest exergy efficiency due to high entropy generation resulting from separation process taking place in the column. These involve momentum loss due to pressure driving force, thermal loss and mass transfer resulting from temperature driving force and mixing of fluids, respectively in the column.

The atmospheric distillation tower and the pre-flash drum have high irreversibility of 314,670.11 and 195,863.10 kW, respectively. The contributions of atmospheric tower and the pre-flash drum tower to the total irreversibility occurring in the process are 55.43 and 34.5%, respectively. The essence of pre-flash is to reduce the heat load needed in the distillation column for separation process. The heated crude, on entering the pre-flash drum, is separated into crude liquid at the bottom and crude gas at the top as a result of sudden pressure drop in the pre-flash drum. Since the entropy is a function of temperature and pressure, the high irreversibility in the pre-flash unit is mainly due to exergy losses resulting from entropy generation following heated crude liquid flashing into crude gas. While the exergy loss in the distillation column is due to entropy generation resulting from temperature variation and pressure drop accompanies separation operation taking place in the distillation column.

4.1.4. Effect of number of trays on overall exergy efficiency and irreversibility

The capital cost of crude distillation column is highly dependent on the number of trays making up the column. Since the distillation column is the major unit that determines the overall exergy efficiency of the crude distillation process, the effect of increasing number of trays on the overall exergy efficiency and irreversibility was investigated and the result presented in Figure 18. It could be observed from the figure that the overall exergy efficiency increases with increasing number of trays while the exergy loss (irreversibility) was reducing. This is because increasing the number of trays results in better separation. The reflux and liquid flow to the bottom of the column are reduced. The pressure drop within the column increased, therefore the entropy generation due to liquid mixing resulting from separation within the column also reduced. Hence, overall exergy efficiency increases while the overall irreversibility reduces with increasing number of trays.

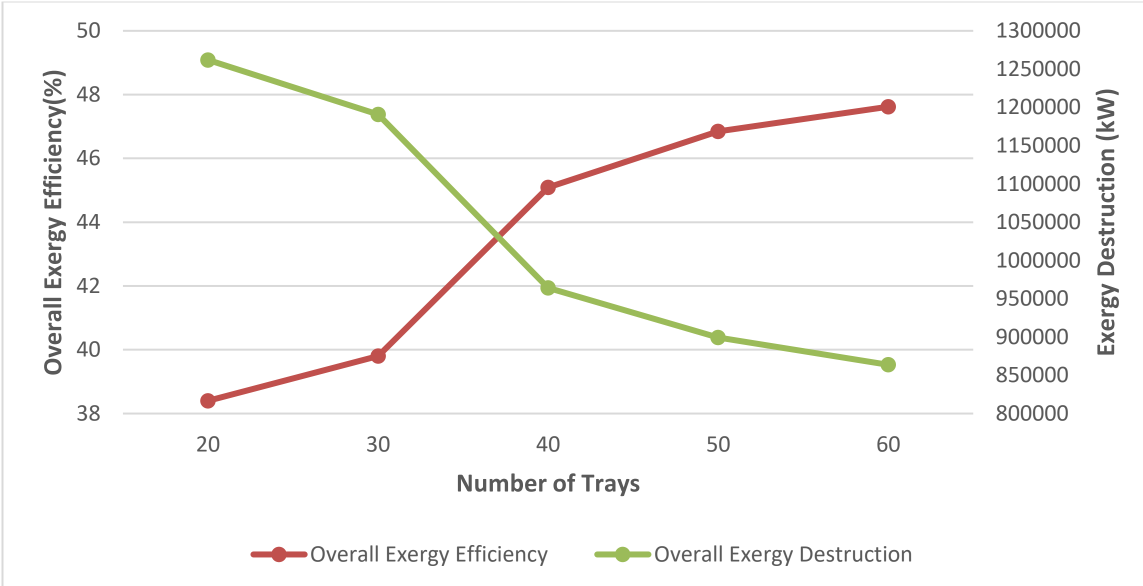


Figure 18: Effect of number of trays on overall exergy efficiency and irreversibility

4.2 Pinch Analysis

Pinch analysis has been carried out in an Aspen Energy Analyzer (AEA) at ΔTmin of 10°C. Two cases have been developed from process streams of heat exchanger network i.e. base and optimization case. The base case has 20 heat exchangers in the preheat train network to increase the inlet temperature of crude to the distillation column ultimately. In contrast, the optimization case is developed after optimization analysis in the Aspen energy analyzer (AEA) to decrease the number of heat exchangers to 18, keeping the temperature recovery the same and conserved. This has been done by different looping and optimum network design of heat exchangers network.

4.2.1. Composite Curve

Composite curves are T/H diagrams (temperature/heat diagrams) used to visualize cold and hot streams and potential heat transfer between them. This curve shows heating utility requirements of 84890000 kJ/h, and that of cold utilities are 2251000 kJ/h for maximum energy requirements and maximum heat transfer between hot and cold streams.

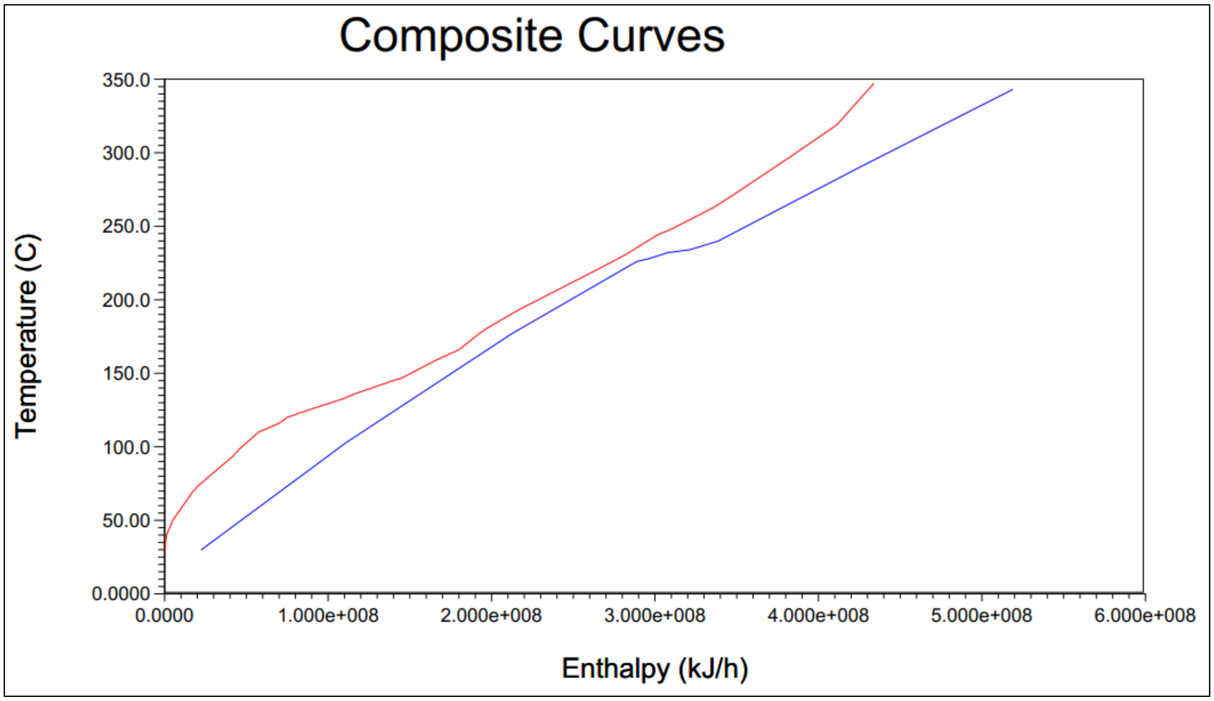


Figure 19: Composite curve of heat exchanger network

4.2.2. Grand Composite Curve

The grand composite curve is a graphical representation of the excess heat available to a process within each temperature interval. In intervals where a net heat surplus exists, we cascade that heat to lower temperature intervals. This T-H diagram in figure 20 shows the regions of the network where there is a need for heating and cooling utilities.

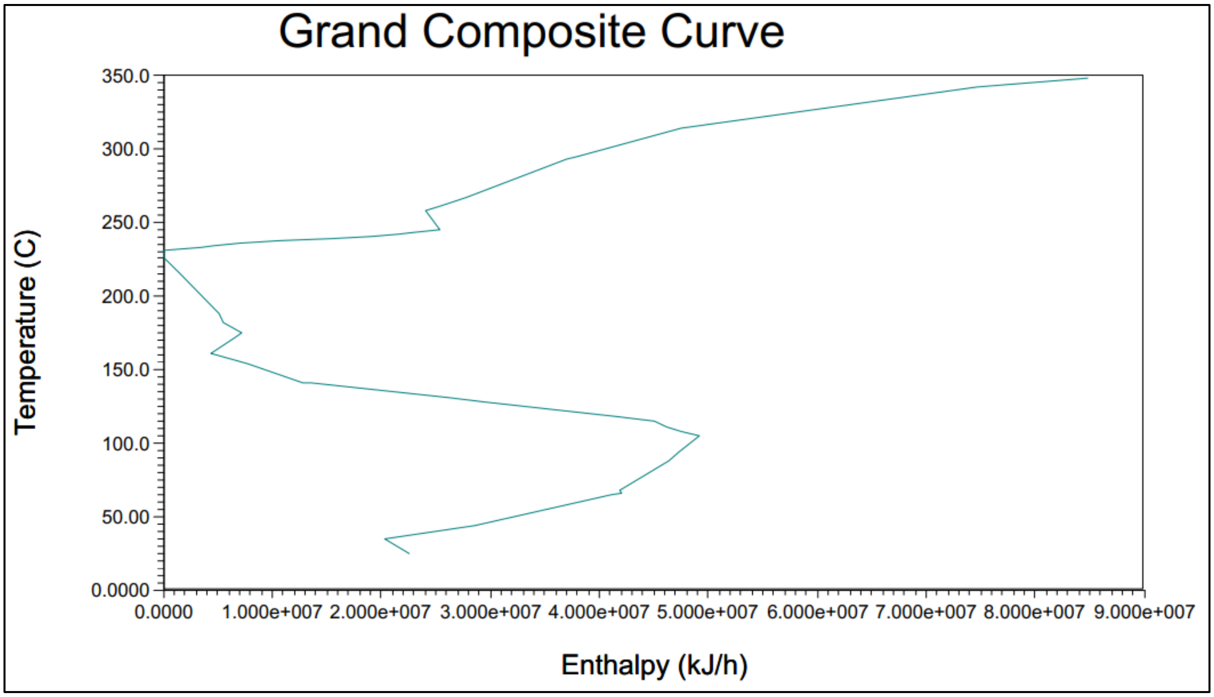


Figure 20: Grand Composite Curve of heat exchanger network

4.2.3. Heat Exchanger Network (HEN) design – base case

The entire heat exchanger network design maximizes the amount of energy recovered from the process heat by utilizing as many heat transfer units—heat exchangers, coolers, and heaters/furnaces—as needed. The afore mentioned utility targets were achieved by this network (base case), as shown in Figure 21. Twenty heat exchangers, including heaters/furnaces, coolers and heat exchangers, are used to achieve the targets.

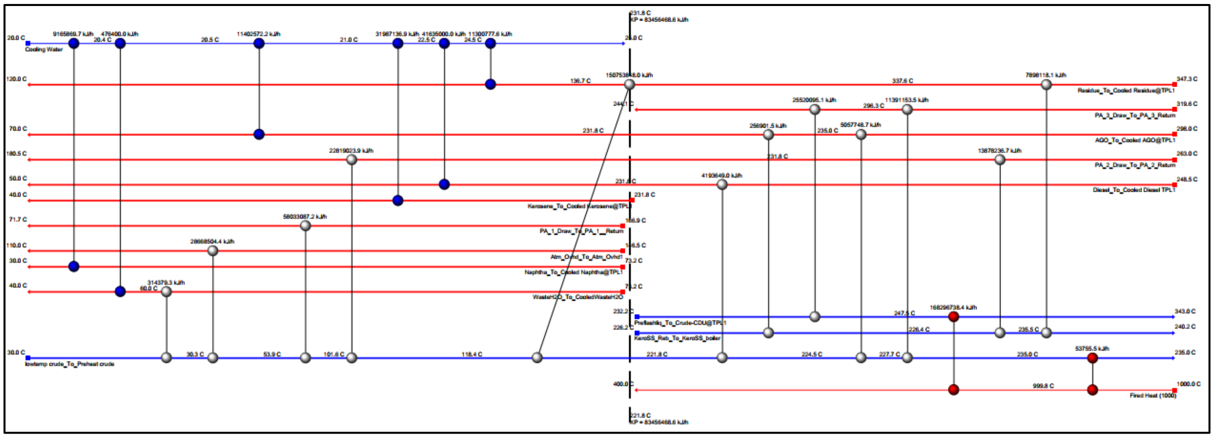
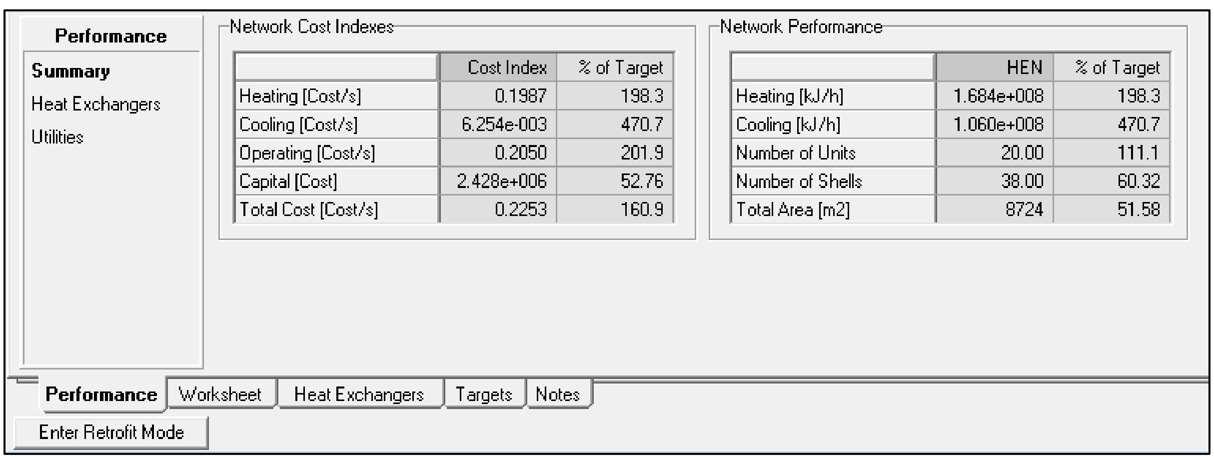


Figure 21: maximum energy recovery base case network

4.2.4. *Base case performance*

Base case heat exchanger network design is used to fulfill the heating and cooling requirements as well as keep the cost index minimum. The software results for the base case are shown in figure 22 below;



Performance

Summary

Heat Exchangers

Utilities

Network Cost Indexes

	Cost Index	% of Target
Heating [Cost/s]	0.1987	198.3
Cooling [Cost/s]	6.254e-003	470.7
Operating [Cost/s]	0.2050	201.9
Capital [Cost]	2.428e+006	52.76
Total Cost [Cost/s]	0.2253	160.9

Network Performance

	HEN	% of Target
Heating [kJ/h]	1.684e+008	198.3
Cooling [kJ/h]	1.060e+008	470.7
Number of Units	20.00	111.1
Number of Shells	38.00	60.32
Total Area [m2]	8724	51.58

Performance | Worksheet | Heat Exchangers | Targets | Notes

Enter Retrofit Mode

Figure 22: Performance data of base case network

4.2.5. *Heat exchangers & utilities data – base case*

The following table 7 shows the heat exchanger summary data of the base case heat exchanger network, which includes the required cost index, area, number of shells and heat load of exchanger calculated by an aspen energy analyzer (AEA). In table 8, utility stream data results are shown for the base case network.

Table 7: Base case heat exchangers

Sr#	Heat Exchanger	Cost Index	Area (m^2)	Shells	Heat Load (kJ/hr)
1	E-196	222228.7	900.2	2	25520095.0
2	E-177	60625.6	150.0	2	41635000
3	E-222	27745.3	48.1	1	11300777.5
4	E-137	113526.8	367.0	2	13878236.7
5	E-203	23689.07	34.8	1	7898118.1
6	E-221	592968.3	2419.1	6	150753848
7	E-158	11526.4	2.24	1	476400
8	E-228	46980.7	101.3	2	5057748.7
9	E-198	56449.3	160.2	1	4193649.0
10	E-231	10224.0	0.2036	1	53755.5
11	E-213	55425.4	155.8	1	22819023.9
12	E-179	51988.3	118.7	2	31987136.9
13	E-162	113121.5	434.3	1	28668504.3
14	E-157	14234.0	8.02	1	314379.3
15	E-227	39249.2	89.9	1	11391153.5
16	E-217	28162.5	49.5	1	11402572.1
17	E-110	15315.1	10.6	1	256901.5
18	E-163	43964.2	108.32	1	9165869.7
19	E-164	515193.5	2022.65	6	58033087.16
20	E-226	384964.1	1542.0	4	168296738.4

Total	**2427583**	**8723.8**	**38**	**603102995.7**

Table 8: Base case utilities

Utility	**Cost index**	**Heat Load (kJ/hr)**	**% of Target**
Cooling Water	0.00625	106,000,000	470.73
Fired Heat (1000)	0.199	168,400,000	198.31

4.2.6. Heat Exchanger Network (HEN) design – optimized case

This stage aims to decrease the number of pathways and loops rather than completely remove them. In order to reduce capital costs, this stage entails eliminating heat exchangers by permitting a minor energy penalty, which lowers the number of units in the network. The goal is to design a balanced heat exchanger network that minimizes capital costs associated with the heat exchanger areas while maintaining the topology that maximizes energy transfer between the streams. The final optimized network is presented in Figure 23.

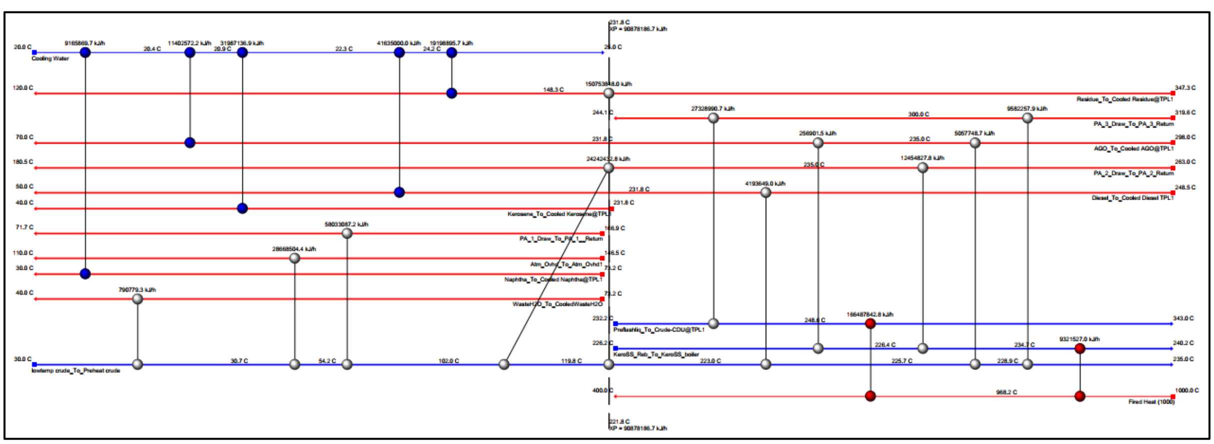


Figure 23: Optimized and final recovery network

4.2.7. Optimized case performance

Optimized case heat exchanger network design is developed via aspen energy analyzer software with two fewer heat exchangers than the base case to fulfill the heating and cooling requirements. The software results for the optimized case are shown in figure 24 below;

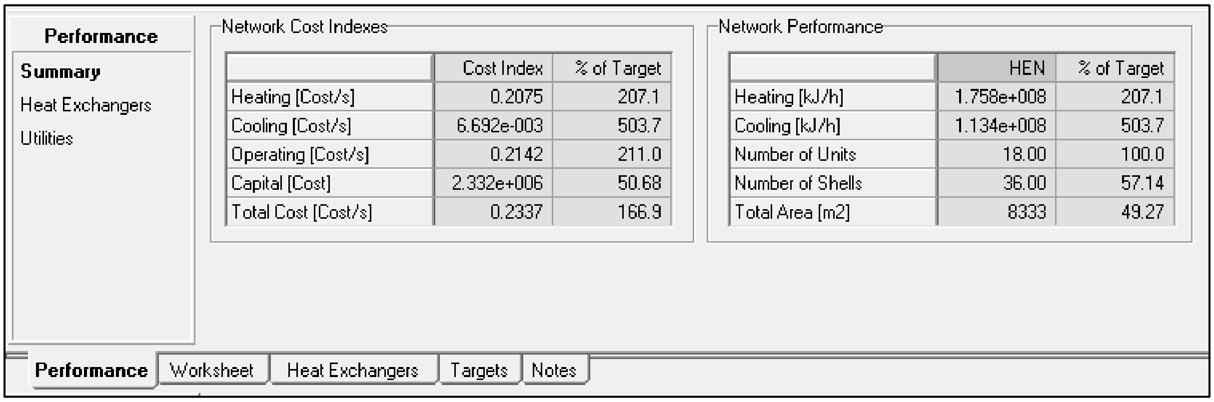


Figure 24: Performance data of optimized case network

4.2.8. Heat exchangers & utilities data – optimized case

The following table 9 shows the heat exchanger summary data of the optimized case heat exchanger network, which includes the required cost index, area, number of shells and heat load of exchanger calculated by Aspen Energy Analyzer (AEA). In table 10, utility stream data results are shown for the base case network.

Table 9: Optimized case heat exchangers

SN	Heat Exchanger	Cost Index	Area (m^2)	Shells	Heat Load (kJ/hr)
1	E-196	228489.487	933.6158122	2	27328990.69
2	E-177	60470.78018	149.5133733	2	41635000
3	E-222	35996.50614	77.58562198	1	19198895.66
4	E-137	90646.6748	268.6043855	2	12454827.84
5	E-221	504682.2199	1970.187594	6	150753848

6	E-228	48606.44518	106.9559095	2	5057748.738
7	E-198	59958.25911	175.5483033	1	4193649.005
8	E-213	56677.08628	161.2564851	1	24242432.76
9	E-179	51907.91968	118.5086759	2	31987136.91
10	E-162	113508.0936	436.3705204	1	28668504.36
11	E-157	23011.59272	32.66258795	1	790779.3128
12	E-227	35097.29773	74.24568521	1	9582257.916
13	E-233	23175.56253	33.17790616	1	9321526.973
14	E-217	28151.60834	49.52008783	1	11402572.17
15	E-110	15315.19082	10.66684766	1	256901.5232
16	E-163	43945.20759	108.2955478	1	9165869.712
17	E-164	520119.3139	2047.337363	6	58033087.16
18	E-226	392107.6621	1578.900278	4	166487842.8
	Total	**2331866.908**	**8332.952986**	**36**	**610561871.5**

Table 10: Optimized case utilities

Utility	**Cost index**	**Heat Load (kJ/hr)**	**% of Target**
Cooling Water	0.006692	113,400,000	503.7
Fired Heat (1000)	0.2075	175,800,000	207.1

From the above results, heat exchanger data in m^2 is reduced, which will ultimately reduce the capital cost of the optimized case. On the other hand, utility requirements in the optimized case will increase but that cost will not affect the profitability of crude distillation unit and can be recovered.

4.2.9. Comparison of the base case and optimized designs

There are many loops and pathways in the base MER network; eliminating some or all of these loops and paths is not the goal of the optimized step for this design. In comparison to the simulation base case design, this network contains two less heat transfer units. Table 11 presents the cost implications of the two considered networks. The optimized case of the process reduces the capital cost from $2427582.99 to $2331866.91, as explained below prior to the optimization results of the process using the pinch technology. Around 3.94% of capital cost was saved by optimizing the heat exchanger network of the crude distillation process.

Table 11: Cost comparison of base and optimized case based on economic

Exchanger Network Design	**Total cost index (cost/s)**	**Area (m^2)**	**Units**	**Shells**	**Capital Cost index ($)**	**Heating (kJ/hr)**	**Cooling (kJ/hr)**	**Operating Cost index (cost/s)**
Simulation base design	0.225	8724	20	38	2427583	168350494	105967756	0.205
Optimized design	0.234	8333	18	36	2331867	175809370	113389475	0.214
Saving Capital ($)	95716.08							
Percentage Saving Capital (%)	3.94							

4.3 Economic Analysis

The economic analysis is done to estimate the total annual cost, capital cost, utility cost and operating cost of the process. The cost analysis also presents the total product of refined crude. For this analysis, 365 operation days are taken as the basis for a plant in a year that makes 8765 hours of total working period. Moreover, Aspen Process Economic Analyzer (APEA) was implemented for the evaluation of the cost of utilities and equipment obtained from the APEA software environment. The utility cost was estimated to be 6,733,850 USD/year, While the total equipment cost sums to 9,165,400 USD/year.

4.3.1. Total Direct and indirect cost

Polanco Martínez et al. [27] analyses the relationship between oil time series in the time-scale domain, such as the prices that come from the complex system formed by crude oil and petroleum products. The prices of raw crude and gasoline products are taken from that research using daily market prices from 2006 to 2017, as shown in Table 12.

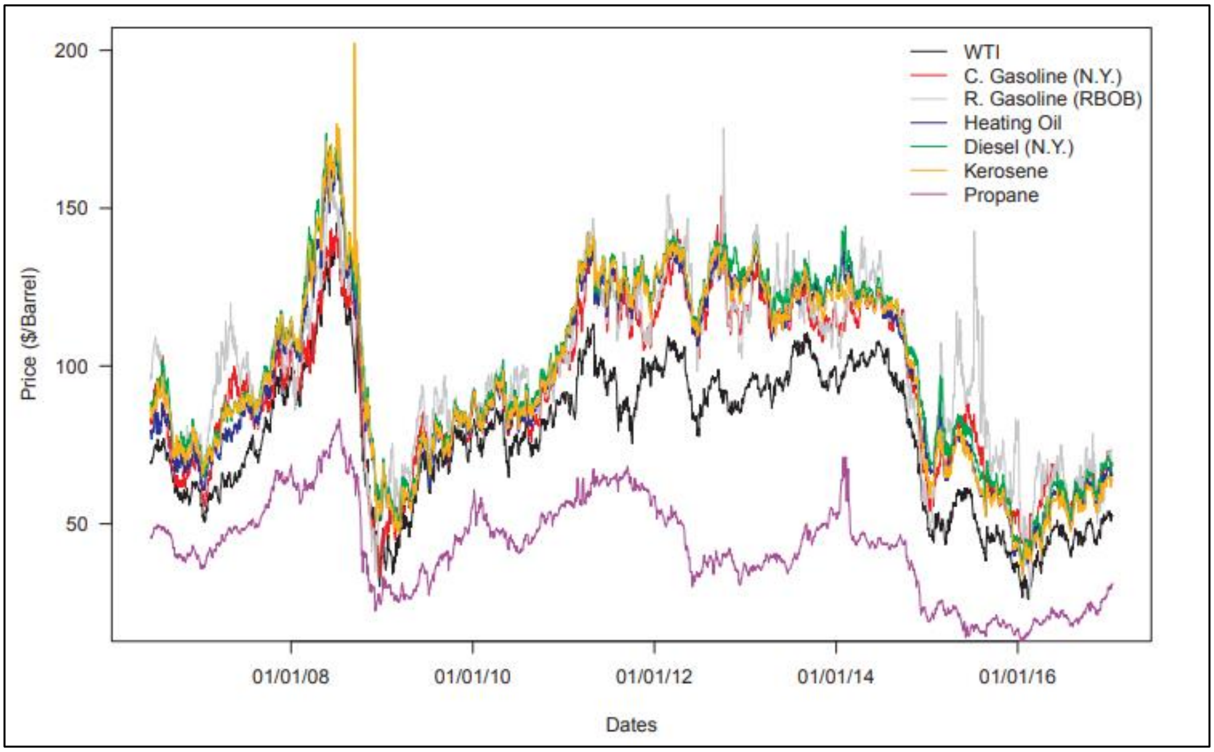


Figure 25: crude oil prices and prices of six refined products for 14/06/2006–17/01/2017 (daily data) [27]

Table 12: Cost of Raw Material and Product

Material Stream	**Cost**
Raw Crude (low-temperature crude)	$77.8/barrel
Naphtha (Distillate)	$93.18/barrel

Kerosene	$97.04/barrel
Diesel	$99.68/barrel
AGO	$95.67/barrel
Residue	$45.7/barrel

For the estimation of total direct cost, it was required to find the cost of every piece of equipment being used in the process of crude distillation unit. The Aspen process economic analyzer (APEA) environment is used for the HYSYS simulation to find out the cost of every piece of equipment used in the crude distillation flowsheet, as shown in Table 13.

Table 13: Equipment cost summary from Aspen Economic Analyzer

Area Name	**Component Name**	**Total Direct Cost**
Miscellaneous Flowsheet Area	PreFlash	548600
Miscellaneous Flowsheet Area	Mixer	1.14E+06
Miscellaneous Flowsheet Area	HEN-1	961200
Miscellaneous Flowsheet Area	HEN-2	398200
Miscellaneous Flowsheet Area	Furnace	3.55E+06
Preheat Train	E-106_@Preheat Train	341700
Preheat Train	E-117_@Preheat Train	101300
Preheat Train	E-101_@Preheat Train	155400
Preheat Train	E-119_@Preheat Train	171100
Preheat Train	E-104_@Preheat Train	100100
Preheat Train	KeroSS_Reboiler_@Preheat Train	237200

Preheat Train	E-108_@Preheat Train	279400
Preheat Train	FH-101-D_@Preheat Train	1.19E+06
Preheat Train	E-102_@Preheat Train	184300
Preheat Train	E-109_@Preheat Train	112600
Preheat Train	E-110_@Preheat Train	105800
Preheat Train	E-103_@Preheat Train	187400
Preheat Train	E-118_@Preheat Train	192600
Preheat Train	E-105_@Preheat Train	133000
Preheat Train	E-116_@Preheat Train	377700
Preheat Train	E-111_@Preheat Train	115100
Preheat Train	E-120_@Preheat Train	255400
Preheat Train	E-113_@Preheat Train	108200
Preheat Train	E-115_@Preheat Train	174700
Preheat Train	E-114_@Preheat Train	297700
Preheat Train	E-100_@Preheat Train	83100
Preheat Train	E-107_@Preheat Train	1.26E+06
Preheat Train	E-112_@Preheat Train	641500
Atmos Tower	DieselSS_@Atmos Tower	383500
Atmos Tower	KeroSS_Reboiler @Atmos Tower-reb	88700
Atmos Tower	KeroSS_Reboiler_@Atmos Tower-reb	147700
Atmos Tower	PA_1_Cooler_@Atmos Tower	266200
Atmos Tower	PA_2_Cooler_@Atmos Tower	134600
Atmos Tower	Condenser_@Atmos Tower-cond	179600

Atmos Tower	Condenser_@Atmos Tower-cond acc	271400
Atmos Tower	Condenser_@Atmos Tower-reflux pump	98500
Atmos Tower	AGOSS_@Atmos Tower	333200
Atmos Tower	KeroSS_@Atmos Tower	334700
Atmos Tower	PA_3_Cooler_@Atmos Tower	220700
Atmos Tower	Main TS_@Atmos Tower	2.75E+06
Total		**9165400**

The total capital investment of the processing plant is the sum of fixed capital investment (FCI) and working capital investment (WCI). The FCI is the sum of direct and indirect costs at the expense of different operations that are carried out in the plant. The table below presents the operations that are involved in estimating the direct cost, which are percentages of purchased equipment cost and the estimation of indirect cost, which are the percentages of total direct cost. FCI is estimated to be **$ 29,420,934,** and WCI which is 15% of FCI, calculates to be **$ 4,413,140**. Hence the total capital investment that is calculated after adding FCI and WCI is $ **33,834,074.**

Table 14: Total Direct Cost Estimation

Cost Type	**Value**	**Cost**
Purchased Equipment	100%	9165400
Installation	40%	3666160
Instrumentation & Control	15%	1374810
Piping	50%	4582700
Electricity	10%	916540
Building	15%	1374810

Land	4%	366616
Service Facility	40%	3666160
Yard Improvement	10%	916540
Insulation Cost	8%	733232
Total Direct Cost ($/yr)		**26762968**

Table 15: Total Indirect Cost Evaluation

Cost Type	**Value**	**Cost**
Engineering & Supervision	8%	733232
Contractor Fee	3%	274962
Construction Expenses	10%	916540
Contingencies	8%	733232
Total Indirect Cost ($/yr)		**2657966**

4.3.2. Production cost

Moreover, the total production cost will be estimated based on the sum of variable costs, fixed charges and overhead charges. The utility cost of the plant, taken from the Aspen HYSYS environment in energy analysis, is 6,733,850 USD/yr.

Maintenance cost = 7% of FCI

Maintenance cost = $ 2059465.38

Miscellaneous Material = $ 205946.54 (10% of maintenance cost)

Raw Material Cost:

Flow rate of crude = 519 tonne/hr

For 365 days of operating time = 1,878,520,000 $/yr

Total cost of raw material = 1,878,520,000 $/yr

Variable cost = raw material cost + miscellaneous cost + utilities cost

Variable cost = $ 1,885,459,796.54 $/yr

Fixed Operating Cost:

Direct production cost = variable cost + fixed cost

Fixed cost for operating on the plant: the estimation of fixed cost is shown in Table 16 below.

Table 16: Direct and Annual Production Cost

Cost Type	Typical values	Estimated Values
Variable Costs = A		
Raw materials	Calculated	1,878,520,000
Miscellaneous Materials	10% of Maintenance cost	205,946.54
Utilities	From Aspen Economizer	6,733,850
Sub Total A		**1,878,520,000**
Fixed Costs = B		
Maintenance	5-10% of FCI	2,059,465.38
Operating Labour	From manning estimates	1,010,000
Laboratory Cost	20-23% of Operating Labour	232,300
Supervision	20% of Operating Labour	202,000
Plant overhead	50% of Operating Labour	505,000
Capital Charges	10% of FCI	2,942,093
Insurance	1% of FCI	294,209.34
Local taxes	2% of FCI	588,418.68

Royalties	1% of FCI	294,209.34
Sub Total B		**8127965**
Direct Production Cost = A+B		**1,893,587,493**
Sales expenses = C		
General Overheads	20-23% of Direct Production Cost	568,076,247.90
Research and Development		
Sub Total C		**568,076,247.90**
Annual Production Cost = A+B+C		**2,461,663,740.58**
Production Cost= $\frac{\textbf{Annual production cost}}{\textbf{Annual production rate}}$		**0.001848 \$/tonne or 6.734 \$/barrel or 0.002359 \$/m3**

Production cost = Annual production cost / Annual production rate

Annual Production rate = 519 tonne/hr or 4166.67 barrel/hr

Total Production Cost = 2,461,663,740.58 \$/yr

Refined crude Production Rate = 519 * 8766 = 4,549,554 tonne/yr or 36,525,000 barrel/yr

Production Cost = Total production cost / Refined crude production rate

Production cost =0.001848 \$/tonne or 6.734 \$/barrel or 0.002359 \$/m^3

4.3.3. Profitability Analysis

Selling Price:

Selling price of product = 431.27 \$/barrel

Profit:

Profit = Selling price - production cost

Profit = 424.53 \$/barrel

Total Production per year = 36,525,000 barrel/year

Profit per year = 15,506,170,515 \$/year

Total Income:

Selling Price = 431.27 \$/kg

Total Production per year = 36,525,000 barrel/year

Total Income = 15,752,136,750 \$/year

Gross Profit:

Gross Profit = Total Income - Total Production Cost

Gross Profit = 15,752,136,750 - 2,461,663,740.58

Gross Profit =13,290,473,009.42 \$/year

Net Profit:

Let the tax rate be 30%

Taxes = 0.3*Gross Profit =3,987,141,902.83 \$/year

4.3.4. Depreciation

Assume that the Fixed Capital Investment depreciates by straight-line method for 10 years. Assuming 1 % Salvage value at the end of plant life

$$Depreciation\ = \frac{V - VS}{n}$$

V = F.C.I =29,420,934 \$

VS = 0.05 × F.C. I = 1,471,046.7 $

n = Number of Years = 10 Years

$Depreciation\ = \frac{29{,}420{,}934 - 1{,}471{,}046.7}{10} = 2{,}794{,}988.73$ $/year

4.3.5. Payback period calculation

Net profit = gross profit – depreciation

Net profit = 13,287,678,020.69 $/year

Rate of Return

$ROR\ = \frac{Net\ Profit}{Total\ Income} = \frac{13{,}287{,}678{,}020.69\ \$/\text{year}}{15{,}752{,}136{,}750\ \$/\text{year}}$ = 0.8435

Rate of return = 84.35%

Payback Period:

$$Payback\ Period\ = \ 1/rate\ of\ return\ Payback\ period\ \ = \ 1.18\ year$$

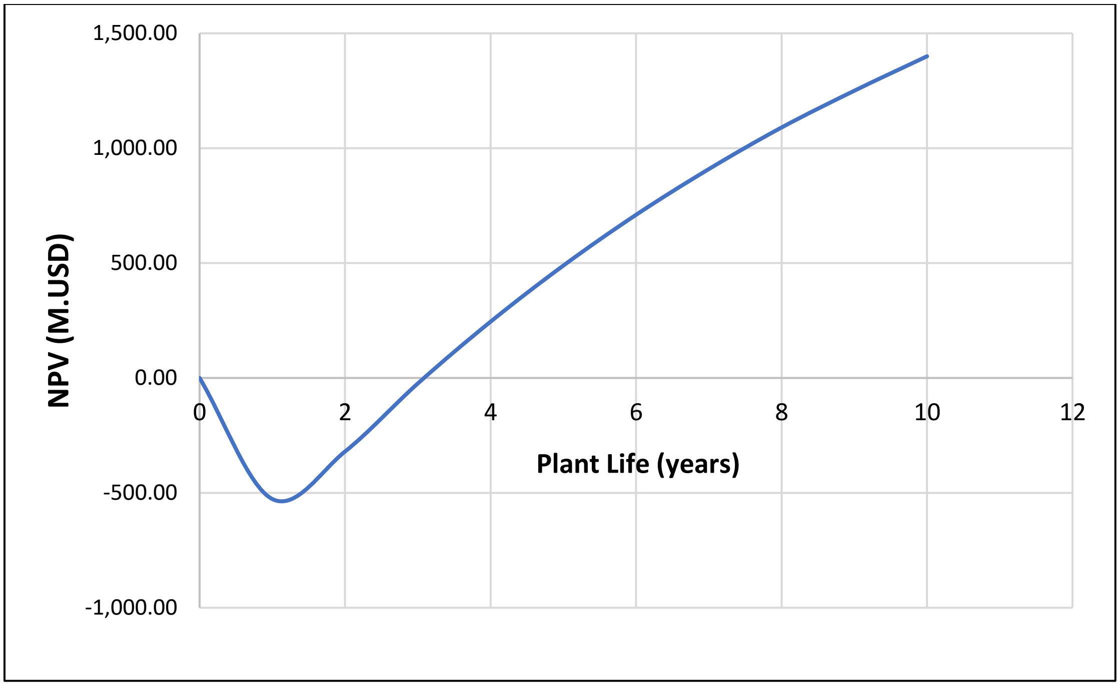


Figure 26: Net Present Value versus Plant life for the process

4.3.6. Comparative analysis of base and optimized case based on economic analysis

Table 17 below presents the optimized percentages in order to measure the impact of optimization and its viability for in-the-moment enhancements in crude refinery plants.

Table 17: A comparative analysis based on economic analysis between the base case and optimized case

Economic Data	Typical Values	Before Optimization	After Optimization	Percentage Optimization (%)
Total Equipment Cost	ASPEN	9165400	9069683.92	3.94
Fixed Capital Investment (FCI)	Direct cost + Indirect cost	29,420,934	29113685.38	
Working Capital	15% of FCI	4,413,140	4367052.81	

Investment (WCI)				
Total Capital Investment (TCI)	FCI+WCI	33,834,074	33480738.19	
Utility Cost	ASPEN	6,733,850	6,759,637.92	
Variable Cost	Raw Material + Utility cost + Miscellaneous	1,885,459,796.54	1,885,483,434	
Fixed Operating Cost (FOC)	Calculated	8,127,696.14	8,013,873.93	
Direct Production Cost (DPC)	Variable Cost+FOC	1,893,587,492.68	1,893,497,307.65	
Overhead Charges	30% of DPC	568,076,247.80	568,049,192.29	
Total Production Cost (TPC)	DPC+Overhead Charges	2,461,663,740.58	2,461,546,499.94	
Gross Profit	Total Income - TPC	13,290,473,009.42	13,290,590,250.06	
Net Profit	Gross Profit-Depreciation	13,287,678,020.69	13,287,824,449.95	
Rate of Return (ROR)	Net Profit/Total Income	0.8435	0.8436	
Payback Period	1/ROR	1.1855	1.1854	

4.3.7. Discussion of the results

Exergy Analysis is employed to identify energy inefficiencies within the CDU. Initially, the steady-state exergy analysis was carried out using an Aspen HYSYS model to quantify the exergy destructions and exergy efficiencies of all the individual heat exchangers and the overall Heat Exchanger Network. The overall exergy destruction and exergy efficiency of the HEN was 17611.2 kW and 63.34%, respectively. Exergy Analysis reveals that the energy efficiency can be improved for some of the heat exchangers in the network as they contribute 56 % to the overall exergy destruction of HEN. Among the heat exchangers, E-

113 was least efficient, followed by E-117, E-103, E-102 and E-110 with exergy efficiency of 6.679768%, 8.388046%, 11.29506%, 11.68023% and 13.02756% respectively.

In comparison, E-108 has the highest exergy efficiency of 94.34035%. Figure 16 shows the visual presentation of exergy efficiencies of individual heat exchangers belonging to the HEN. Effective heat exchanger design can limit the losses of heat and pressure. The exergy efficiency of heat exchangers can be improved by enhancing the thermal designs of heat exchangers, keeping in mind the techno-economic aspects.

PA, facilitated by Aspen Energy Analyzer (AEA), for pinch analysis of the existing preheat trains of crude distillation unit and the optimized case, is also developed for more energy integrated and efficient design of crude preheat train and distillation unit. The minimum temperature approach was optimized to obtain the optimum ΔT_{min} of 10 ℃ for the minimum total annualized cost for the energy analysis. The final heat exchanger network design, based on this optimum ΔT_{min}, is also presented along with its composite curve, the grand composite curve. With the analysis of base case CDU, an improved heat exchanger network (HEN) was obtained. Twenty heat exchangers with a surface area of 8723.8 m^2 were used to obtain a minimum annual capital cost (ACC) of $4,601,000/yr, annual operating cost (AOC) of $3,202,731/yr and total annualized cost (TAC) of $ 4, 417, 688/yr. The optimized case achieved the same energy requirements as of base case with two fewer heat exchangers, reducing the overall capital cost of the crude distillation unit by $ 95716.08 despite the increase in heating and cooling utility costs. Finally, a payback period of 1.18 years has been calculated through an economic analysis for being a crude distillation unit operational.

Moreover, the economic analysis conducted as part of this study yields a profound understanding of the economic landscape of CDU. The total equipment cost, comprised of heat exchangers and a crude distillation unit, provides an exact grasp of the process's capital-intensive features. This fits in with the overall capital expenditure, which considers costs for project management, engineering, and construction in addition to equipment. The precise computation of direct production costs, which include labor, maintenance, overheads, and raw materials, provides insight into the economic impact of daily operations. By comparing revenues against total costs, the calculation of net profit also provides a strong indicator of the economically feasible nature of refined crude production.

Alongside this, the rate of return turns into an essential indicator that measures the startup's profitability and desirability to possible investors. The estimated payback period, which represents the amount of time required for the initial capital expenditure to be recovered through net profits, completes the economic story and offers a realistic assessment of the project's resilience and sustainability in terms of economics. When every factor is considered, the outcomes from the economic study provide an overview of the feasibility of the production of crude refining, providing stakeholders with the knowledge they need to make well-informed decisions and develop strategic plans. In addition, a comparison of the process before and after optimization is made in order to determine the significance of pinch analysis.

4.4 Limitations of the study

Assumptions in Modeliing: The correctness of the study is highly dependent on the assumptions established during modeling in Aspen HYSYS, AEA, and APEA. The validity of the results may be compromised if there are any disparities between the assumptions made and the actual situations in the real world.

Data availability is a crucial factor in ensuring the correctness of economic research. The reliability and accessibility of data related to equipment costs, operating expenses, and market pricing play a significant role in this respect. However, it is important to note that such data may not always be readily accessible or up-to-date.

Scope: The study specifically concentrates on the crude preheat train within the CDU, possibly overlooking other portions of the refinery that could also be optimized. By broadening the scope to encompass more units or processes, one can gain a more comprehensive insight into the potential for energy integration.

CHAPTER 5: CONCLUSION

5.1 Conclusion

Exergy Analysis is employed to identify energy inefficiencies within the CDU. Initially, the steady-state exergy analysis was carried out using an Aspen HYSYS model to quantify the exergy destructions and exergy efficiencies of all the individual heat exchangers and the overall Heat Exchanger Network. The overall exergy destruction and exergy efficiency of the HEN was 17611.21 kW and 63.34%, respectively. The exergy efficiency of heat exchangers can be improved by enhancing the thermal designs of heat exchangers, keeping in mind the techno-economic aspects. 40 trays are found to be optimal for the efficient design of crude distillation column while keeping exergy destruction minimum. Pinch analysis was successfully applied to study the energy and economic implications of the Crude refining unit using Aspen Energy Analyzer Software. The work relied on real process data which were extracted using Aspen HYSYS for subsequent use in HEN design in Aspen Energy Analyzer. The pinch analysis of the process shows that although the cooling and heating loads were not the same, a restriction was placed on heat exchange by the fewer cold process streams than hot process streams. The HEN uses twenty heat exchangers for the MER base case, while the optimized uses eighteen heat exchanger units, most of which are coolers, to remove heat from many heat sources that outweigh the heat sinks. The optimized case achieved the same energy requirements as of base case with two fewer heat exchangers, reducing the overall capital cost of the crude distillation unit by $ 95716.08 despite the increase in heating and cooling utility costs. Finally, a payback period of 1.18 years has been calculated through an economic analysis for being a crude distillation unit operational. It can be seen from this study that by developing an energy-integrated system using pinch technology as a tool, a large amount of energy can be saved.

5.2 Future Work

Dynamic Simulation: Performing dynamic simulations to consider temporary conditions and operational variations can offer a more accurate evaluation of system efficiency and energy consumption over some time.

Integration of Renewable Energy: An exploration into the incorporation of renewable energy sources, such as solar or waste heat recovery, into the energy systems of the refinery, might significantly improve sustainability and minimize the refinery's environmental footprint.

Advanced Control Techniques: By investigating sophisticated control techniques, such as model predictive control (MPC) or fuzzy logic control, we can enhance the performance of the HEN in real time. This optimization aims to maximize energy efficiency while ensuring that process restrictions and needs are met.

Performing a lifecycle assessment to analyze the environmental effects linked to various optimization strategies might aid in determining the best environmentally friendly solutions for the refinery's operations.

REFERENCES

[1] S. Paramonova, P. Thollander, M. J. R. Ottosson, and S. E. Reviews, "Quantifying the extended energy efficiency gap-evidence from Swedish electricity-intensive industries," vol. 51, pp. 472-483, 2015.

[2] J. Fajardo, C. Negrette, D. Yabrudy, and C. Cardona, "Advanced exergetic analysis of preheat train of a crude oil distillation unit," in *ASME International Mechanical Engineering Congress and Exposition*, 2021, vol. 85642, p. V08BT08A007: American Society of Mechanical Engineers.

[3] C. Yan, L. Lv, A. Eslamimanesh, and W. J. A. T. E. Shen, "Application of retrofitted design and optimization framework based on the exergy analysis to a crude oil distillation plant," vol. 154, pp. 637-649, 2019.

[4] M. A. Durrani, I. Ahmad, M. Kano, and S. J. E. Hasebe, "An artificial intelligence method for energy efficient operation of crude distillation units under uncertain feed composition," vol. 11, no. 11, p. 2993, 2018.

[5] D. A. Kamel, M. A. Gadalla, O. Y. Abdelaziz, M. A. Labib, and F. H. J. E. Ashour, "Temperature driving force (TDF) curves for heat exchanger network retrofit–A case study and implications," vol. 123, pp. 283-295, 2017.

[6] J.-C. Bonhivers, B. Srinivasan, and P. R. J. A. T. E. Stuart, "New analysis method to reduce the industrial energy requirements by heat-exchanger network retrofit: Part 1–Concepts," vol. 119, pp. 659-669, 2017.

[7] F. Bühler, T.-V. Nguyen, and B. J. A. E. Elmegaard, "Energy and exergy analyses of the Danish industry sector," vol. 184, pp. 1447-1459, 2016.

[8] E. S. Dogbe, M. A. Mandegari, and J. F. J. E. Görgens, "Exergetic diagnosis and performance analysis of a typical sugar mill based on Aspen Plus® simulation of the process," vol. 145, pp. 614-625, 2018.

[9] M. Aghbashlo, H. Mobli, S. Rafiee, A. J. R. Madadlou, and S. E. Reviews, "A review on exergy analysis of drying processes and systems," 22, 2013.

[10] S. E. Jorgensen and B. D. Fath, *Encyclopedia of ecology*. Elsevier BV, 2008.

[11] T.-V. Nguyen, M. Voldsund, B. Elmegaard, I. S. Ertesvåg, and S. J. E. Kjelstrup, "On the definition of exergy efficiencies for petroleum systems: Application to offshore oil and gas processing," vol. 73, pp. 264-281, 2014.

[12] M. Mehdizadeh-Fard and F. J. J. o. C. P. Pourfayaz, "Advanced exergy analysis of heat exchanger network in a complex natural gas refinery," vol. 206, pp. 670-687, 2019.

[13] S. Kumar and A. S. J. R. i. E. Mhetre, "Comparative techno-economic evaluation of potential processing schemes for petroleum crude oil distillation," vol. 14, p. 100480, 2022.

[14] A. Costa *et al.*, "Analysis of the environmental and economic impact of fouling in crude preheat trains for petroleum distillation," in *Proceedings of the 9th international conference on heat exchanger fouling and cleaning, Crete Island, Greece*, 2011, pp. 5-10.

[15] M. M. Fard, F. Pourfayaz, A. Kasaeian, M. J. J. o. N. G. S. Mehrpooya, and Engineering, "A practical approach to heat exchanger network design in a complex natural gas refinery," vol. 40, pp. 141-158, 2017.

[16] K. Ajao and H. J. R. Akande, "Energy integration of crude distillation unit using pinch analysis," vol. 1, no. 2, pp. 54-66, 2009.

[17] E. M. Al-Mutairi and B. S. J. A. P. J. o. C. E. Babaqi, "Energy optimization of integrated atmospheric and vacuum crude distillation units in oil refinery with light crude," vol. 9, no. 2, pp. 181-195, 2014.

[18] V. K. Bulasara, R. Uppaluri, and A. K. J. A. t. e. Ghoshal, "Revamp study of crude distillation unit heat exchanger network: Energy integration potential of delayed coking unit free hot streams," vol. 29, no. 11-12, pp. 2271-2279, 2009.

[19] M. A. J. E. Gadalla, "A new graphical method for Pinch Analysis applications: Heat exchanger network retrofit and energy integration," vol. 81, pp. 159-174, 2015.

[20] I. H. Alhajri, M. A. Gadalla, O. Y. Abdelaziz, and F. H. J. C. S. i. T. E. Ashour, "Retrofit of heat exchanger networks by graphical Pinch Analysis–A case study of a crude oil refinery in Kuwait," vol. 26, p. 101030, 2021.

[21] J. Zun-long, D. Qi-wu, L. J. C. E. Min-shan, and T. I. C. P. E. P. Engineering-Biotechnology, "Exergoeconomic analysis of heat exchanger networks for optimum minimum approach temperature," vol. 31, no. 2, pp. 265-269, 2008.

[22] T. Benali, D. Tondeur, and J. N. J. A. T. E. Jaubert, "An improved crude oil atmospheric distillation process for energy integration: Part I: Energy and exergy analyses of the process when a flash is installed in the preheating train," vol. 32, pp. 125-131, 2012.

[23] Z. N. Izyan and M. J. E. Shuhaimi, "Exergy analysis for fuel reduction strategies in crude distillation unit," vol. 66, pp. 891-897, 2014.

[24] T. J. Kotas, *The exergy method of thermal plant analysis*. Paragon Publishing, 2012.

[25] M. Nakaiwa *et al.*, "Internally heat-integrated distillation columns: a review," vol. 81, no. 1, pp. 162-177, 2003.

[26] S. A. Ashrafizadeh, M. Amidpour, and M. J. J. o. c. e. o. J. Abolmashadi, "Exergy analysis of distillation column using concept of driving forces," vol. 46, no. 7, pp. 434-443, 2013.

[27] Polanco Martínez, J. M., Abadie, L. M., & Fernández-Macho, J. (2018). A multi resolution and multivariate analysis of the dynamic relationships between crude oil and petroleum-product prices. Applied Energy, 228, 1550 1560.