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UNIVERSITI TEKNOLOGI PETRONAS

MODELING, SIMULATION AND OPTIMIZATION OF

INDUSTRIAL HEAT EXCHANGER NETWORK

FOR OPTIMAL CLEANING SCHEDULE

by

LOW WAI CHONG

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MODELING, SIMULATION AND OPTIMIZATION OF INDUSTRIAL HEAT EXCHANGER NETWORK FOR OPTIMAL CLEANING SCHEDULE

by

LOW WAI CHONG

A Thesis

Submitted to the Postgraduate Studies Programme

as a Requirement for the Degree of

MASTER OF SCIENCE

CHEMICAL ENGINEERING

UNIVERSITI TEKNOLOGI PETRONAS

BANDAR SERI ISKANDAR

PERAK

SEPTEMBER 2011

DECLARATION OF THESIS

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ACKNOWLEDGEMENTS

My most sincere gratitude goes to my former supervisor, Late Prof. V.R. Radhakrishnan for his guidance and support in developing the ideas and concepts about this research in the initial stage. I also would like to extend my deepest gratitude to my supervisor, Assoc. Prof. M. Ramasamy for his continuous guidance, valuable advices, support and encouragement throughout the period of completing this research.

My sincere thank also goes to all members of Crude Oil Fouling Research Centre (CROFREC) of UTP, especially Totok R. Biyanto, Umesh Deshannavar, M Syamzari and M Zamidi for their cooperation and endless supports to this research.

I would also like to thank Chemical Engineering Department and Postgraduate Office of UTP for providing me an opportunity to pursue my research works.

Last and foremost, I heartily thankful my family members for their patience and support. My special thank to my wife, Cheng Sheau Huoy and my daughter, Low Wen Yan for their love, understanding and supports so that I could concentrate on doing this research.

ABSTRACT

Sustaining the thermal and hydraulic performances of heat exchanger network (HEN) for crude oil preheating is one of the major concerns in refining industry. Virtually, the overall economy of the refineries revolves around the performance of crude preheat train (CPT). Fouling in the heat exchangers deteriorates the thermal performance of the CPT leading to an increase in energy consumption and hence giving rise to economic losses. Normally the energy consumption is compensated by additional fuel gas in the fired heater. Thus, increase of energy consumption causes an increase in carbon dioxide emission and contributes to green house effect. Due to these factors, heat exchanger cleaning is performed on a regular basis either by chemical or mechanical cleanings. The disadvantage of these cleanings is the potential environmental problem through the application, handling, storage and disposal of cleaning effluents. Nevertheless, the loss of production caused by plant downtime for cleaning is often more significant than the cost of cleaning itself, particularly in refineries. Thus, it is essential to optimize the cleaning schedule of heat exchangers in the HEN of CPT.

The present research focuses on the analysis of the effects of fouling on heat transfer performance and optimization of the cleaning schedule for the CPT. The study involves collection and analysis of plant historical operating data from a Malaysian refinery processing sweet crude oils. A simulation model of the CPT comprising 7 shell and tube heat exchangers post desalter with different mechanical designs and physical arrangements was developed under Petro-SIM™ environment to perform the studies.

In the analysis of effects of fouling on heat transfer performance in CPT, the simulation model was integrated with threshold fouling models that are unique to each heat exchanger. The fouling model parameters are estimated from the historical data. The simulation study was performed for 300 days and the analysis indicated that the position of heat exchangers has a dominant role in the heat transfer performance of CPT under fouled conditions. It is observed from this simulation study, fouling of upstream heat exchangers of the CPT will have higher impact to overall heat transfer performance of the CPT. For the downstream heat exchangers, the decline in their heat transfer performances due to fouling can be compensated by the log-mean temperature difference (LMTD) effect, which will reduce or even increase the heat transfer performance of these heat exchangers.

An optimization problem for the cleaning schedule of the CPT was formulated and solved. The optimization problem considered un-recovered energy cost and cleaning cost of the heat exchangers in the objective function. Optimization of the cleaning schedule was illustrated with a case study of simulation over a period of two years. Constant fouling rates that are extracted from the historical data are used to estimate the fouling characteristics of each heat exchanger in the CPT. For the purpose of comparison, a base case was developed based on the assumption that the heat exchangers will be cleaned when the maximum allowable fouling resistance was reached. mixed integer programming approach was used to optimize the cleaning schedule of heat exchangers. An optimized cleaning schedule with significant cost savings has been determined and reported over a period of two years.

ABSTRAK

Mempertahankan prestasi terma dan hidrolik rangkaian alat penukar haba untuk pemanasan awal minyak mentah adalah salah satu perhatian utama dalam industri pernapisan minyak. Hampir keseluruhan ekonomi kilang minyak mentah bergantung kepada prestasi rangkaian alat penukar haba. Kotoran dalam alat penukar haba memburukkan prestasi terma dan menyebabkan peningkatan keperluan bahan bakar dan kerugian ekonomi. Oleh sebab itu, pembersihan rangkaian alat penukar haba sering dilakukan kerana kotoran. Biasanya keperluan tenaga dikompensasi oleh bahan bakar gas tambahan dalam relau. Oleh sebab demikian, peningkatan keperluan terma menyebabkan peningkatan pelepasan karbon dioksida ke udara dan menyumbang pada kesan rumah hijau. Oleh kerana faktor-faktor ini, pembersihan alat penukar haba harus dilakukan secara sistematik dengan menggunakan pembersihan kimia atau pembersihan mekanikal. Kerugian dari pembersihan ini adalah masalah persekitaran yang berpotensi melalui aspek-aspek aplikasi, pengendalian, simpanan, dan pelupusan sampah bahan pembersihan. Namun demikian, kehilangan pengeluaran produk minyak mentah akibat masa untuk membersihkan rangkaian alat penukar haba sering lebih signifikan dari kos pembersihan itu sendiri, khususnya di kilang pernapisan minyak. Jadi, adalah penting dari segi ekonomi untuk mengoptimumkan jadual pembersihan alat penukar haba dalam rangkaian alat penukar haba.

Penyelidikan ini tertumpu pada analisis pengaruh kotoran kepada prestasi perpindahan terma dan optimalisasi jadual pembersihan untuk rangkaian alat penukar haba untuk minyak mentah ke relau yang merupakan salah satu rangkaian alat penukar haba paling umum di kilang pernapisan minyak. Data operasi telah dikumpul dari kilang Malaysia yang memproses minyak mentah manis untuk digunakan sebagai kes kajian. Satu model simulasi rangkaian alat penukar haba untuk minyak mentah ke relau terdiri daripada 7 alat penukar haba yang berbeza dari segi mekanikal desain, susunan fizikal dan kapasiti telah disimulasikan dalam perisian Petro-SIM™ untuk melakukan kajian ini.

Dalam analisis kesan kotoran pada prestasi pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau, model simulasi telah diintegrasikan dengan model kotoran yang unik untuk setiap alat penukar haba. Parameter model kotaran dianggarkan daripada perkiraan data operasi yang terkumpul. Pengajian simulasi telah dilakukan untuk 300 hari. Analisis kotoran menunjukkan bahawa susunan alat penukar haba dalam rangkaian alat penukar haba mempunyai peranan yang dominan dalam prestasi pemindahan terma dari pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau dalam keadaan kotoran. Sebagaimana ditunjukkan dalam analisis simulasi, alat penukar haba pertama pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau mempunyai kesan kotoran yang paling tinggi untuk perpindahan terma pada keseluruhan prestasi pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau . Untuk alat penukar haba yang lain, kesan kotoran pada prestasi pemindahan terma boleh dikompromikan oleh kesan perbezaan suhu log min yang akan melambatkan atau meningkatkan prestasi permindahan terma untuk alat penukar haba.

Oleh sebab itu, satu masalah pengoptimuman untuk jadual pembersihan pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau telah dirumuskan untuk jangka masa dua tahun. Fungsi objektif pengoptimuman mengambilkira kos tenaga dan kos pembersihan alat penukar haba. Optimasi jadual pembersihan diilustrasikan dengan data operasi yang terkumpul dari kilang Malaysia selama tempoh dua tahun. kadar kotoran yang ditentukan daripada data operasi dianggap tidak berubah untuk setiap alat penukar haba dan boleh digunakan untuk menganggarkan kadar kotoran untuk setiap alat penukar haba dalam pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau. Kadar kotoran yang ditentukan mengambilkira jenis minyak mentah yang diproses dalam kilang pernapisan dan alat penukar haba fizikal geometri. Pada awal kajian pengoptimuman, kes asas disimulasikan untuk menentu kos asas operasi yang diperlukan untuk membersihkan alat penukar haba dalam pemindahan terma rangkaian alat penukar haba untuk minyak mentah ke relau apabila mereka mencapai kadar maksimum kotoran. Pendekatan mixed integer programming digunakan untuk mengoptimumkan jadual pembersihan penukar panas berturut-turut di iterasi. Jadual pembersihan telah dioptimumkan dan dilaporkan dengan penjimatan jumlah kos yang signifikan dalam masa dua tahun.

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NOMENCLATURE

Variables	Description	Units
A	Effective heat transfer area	m^2
A_1	Characteristic constant for Eq. 3.29	(-)
A_2	Characteristic constant for Eq. 3.29	(-)
A_3	Characteristic constant for Eq. 3.29	(-)
a	Constant for Eq. 2.12	(-)
C_{bh}	Constant for Eq. 3.13	(-)
C_{cl}	Cost of heat exchanger cleaning per day	RM/day
C_E	Cost of energy per day	RM/GJ
C_p	Mass heat capacity	J/kgC
C_{rb}	Concentration of precursors	kg/m^3
d_i	Tube internal diameter	m
d_o	Tube external diameter	m
E	Activation energy	kJ/mol
F	Correction factor for *LMTD*	(-)
F_c	Pure cross flow area of the shell	m^2
h_i	Tube side heat transfer coefficient	W/m^2K
h_o	Shell side heat transfer coefficient	W/m^2K
J_c	Correction factor for baffle cut and spacing	(-)
J_l	Correction factor for baffle leakage effect	(-)
J_b	Correction factor for bundle bypass flow	(-)
J_s	Correction factor for variable baffle spacing	(-)
J_r	Correction factor for adverse temperature gradient	(-)
K	Watson characterization factor	(-)
k	Thermal conductivity	W/mK
k_w	Thermal conductivity of tube wall material	W/mK
L_{bc}	Exact baffle spacing	m
L_{ic}	Central baffle spacing	m
L_{bo}	Outlet baffle spacing	m

Variables	Description	Units
M	Mass flow rate	kg/h
N_b	Number of baffles	(-)
N_E	Total number of heat exchanger in the CPT	(-)
P	Absolute pressure	Pa
P_c	Critical pressure	Pa
Pr	Prandtl number	(-)
Q	Heat duty	kW
R	Universal gas constant	m^3Pa/mol. K
Re	Reynolds number	(-)
R_f	Thermal fouling resistance	m^2K/kWh
Sc	Schmidt number	(-)
SG	Specific gravity	(-)
S_b	Ratio of the bypass area	(-)
S_m	Overall shell side cross flow area	m^2
S_{sb}	Total tube to baffle leakage area of shell	m^2
S_{tb}	Shell to baffle leakage area within the circle segment occupied by the baffle	m^2
T	Absolute temperature	°C
T_c	Critical temperature	°C
t	Time	day
t_c	Cleaning duration	day
t_p	Total period of operation	day
U	Overall heat transfer coefficient	W/m^2K
V	Molar volume	m^3/mol
v	Velocity	m^2/s
X_1	Intermediate value of Eq. 3.36	(-)
X_2	Intermediate value of Eq. 3.36	(-)
x	Thickness of deposit	m
y	Binary variable indicate the status of heat exchanger	(-)
z	Length of small element region	m
Z_{RA}	Empirically derived constant	(-)

Greek Letters	Description	Units
ε	Effectiveness of a heat exchanger	(-)
ω	Pitzer acentric factor	(-)
α	Deposition constant	m^2K/kWh
γ	Removal constant	$m^2K/kWh/Pa$
β	Constant	(-)
η	Heat transfer efficiency	(-)
ρ	Density	kg/m^3
ψ	Ratio for ψ-P *method*	(-)
μ	Fluid viscosity	Pa-s
τ	Shear stress	N/m^2
λ	Constant for Eq. 3.31	(-)

Subscripts	Description
b	Bulk
c	Cold stream
clean	Clean condition
FBP	Final boiling point
f	Film
h	Hot stream
i	Inlet
IBP	Initial boiling point
min	Minimum
max	Maximum
o	Outlet
s	Surface
t	Time *t*
w	Wall
∞	At infinite time, asymptotic

Abbreviations	Description
ASTM	American Society for Testing and Materials
CDU	Crude Distillation Column
CIT	Coil Inlet Temperature
COT	Coil Outlet Temperature
CPT	Crude Preheat Train
HE	Heat Exchanger
HEN	Heat Exchanger Network
TEMA	Tubular Exchanger Manufacturers Association
VBN	Viscosity Blending Number

CHAPTER 1

INTRODUCTION

Improvement in energy efficiency is of prime importance in process industries globally. A very common approach for improving energy efficiency is the integration of heat using heat exchanger networks (HEN). For example, crude preheat train (CPT) is an important heat recovery unit in refinery operations to provide adequate preheating to the crude oil before it is processed in the atmospheric crude distillation unit (CDU).

Heat exchanger fouling is a well documented problem in process industries. As reported in ESDU [1], fouling of heat exchangers in the CPT is a major problem that costs the industry billions of dollars every year. It is a chronic operational problem that compromises energy recovery and environment welfare [2].

Fouling deteriorates the thermal efficiency of the heat exchangers leading to considerable reduction in heat recovery. It also affects the delicate balance of heat integration. Besides, fouling results in interrupted production, operation downtime and additional maintenance costs that lead to high cost penalties. In many heat exchanger applications, where the fouling potential of a particular fluid stream is not properly recognized or inadequately allowed for the design, frequent cleaning will be required. Unexpected shut down at short notice will have significant effect on production schedules and overall throughput. In some instances, it is possible to bypass the severely fouled heat exchanger while the production is maintained. The loss of heat duty is supplemented by pre-heating equipment downstream of this heat exchanger. The short fall in the energy recovery in the HEN will have to be made up by an increased use of purchased primary energy.

Fouling in crude preheat train is a very complex phenomenon. Fouling mechanisms and rates differ very much among the heat exchangers in the crude preheat train. Fouling rates depend very much on the operational conditions such as the temperatures and velocities of the hot and cold fluids and crude oil types such as paraffinic, naphthenic, etc.

In general, fouling can be caused by several mechanisms [3]. These mechanisms include sedimentation fouling, crystallization fouling, chemical reaction fouling, corrosion fouling and biological fouling. Sedimentation fouling happens when suspended solids in the fluid settle out on the heat transfer surfaces. Crystallization fouling is related to salts in a fluid that crystallize on the surface when it encounters wall temperatures above the saturation point of the dissolved salt. As compared to sedimentation fouling and crystallization fouling that only involves primarily physical changes, chemical reaction fouling produces solid phase at or near the heated surface. For example, carbonaceous deposits or commonly recognized as coke formation due to thermal degradation of one of the process fluid components. Corrosion product fouling is due to fluids that corrode the metal of the heat transfer surface. Biological fouling is common for heat exchange fluid that is based on cooling water source or process streams that contain micro organisms. The organisms will attach to solid surface and grow, and hence reduce the heat transfer.

Online hot melting technology is used to remove the foulant from the surfaces of the heat exchangers in the CPT. Normally diesel is used as the washing agent. This technique boosts performance as it does not only eliminate staff time for cleaning services, but also purchase of chemicals and waste disposal. Nevertheless, this technique cannot clean the foulant of heat exchangers entirely, which means more frequent cleaning cycles of heat exchangers are needed.

Mechanical cleaning of fouled heat exchangers is carried out by isolating the heat exchanger, which results in lost production. The mechanical cleaning procedures are labor-intensive and expensive. Nevertheless, it is able to remove the foulant completely and hence restore the heat transfer performance. Furthermore, the offline period of the mechanical cleaning while heat exchanger is isolated is one third of the duration needed to perform the cleaning of heat exchanger online [3].

Maintenance costs to remove fouling deposits, costs of chemicals and other operating costs of antifouling devices are about 15% of the maintenance costs of process plants [4]. In addition to production losses due to shut down of plants for cleaning of fouled heat exchangers, penalties for not keeping to a deadline and the loss of customers must be considered. Thus, it is essential to have a well-planned cleaning cycles of heat exchangers in the HEN to minimize the maintenance cost.

In general, the fouling cannot be avoided, but can be mitigated. The mitigation approaches include: i) additional heat transfer allowance by increasing heat transfer area during design stage [5]; ii) application of continuous helical baffles instead of segmental baffles in conventional shell and tube heat exchangers for process fluids that have high fouling tendency at the shell side [6]; iii) addition of standby heat exchangers to avoid sudden shut down of the plant operation due to fouling; iv) corrosion resistant materials for tubes where the corrosion induced fouling is predominant [7]; v) coating of tube surfaces to modify the surface properties [8]; vi) use of specially designed tubes that provides additional heat transfer coefficient without any increase in the pressure drops (Bouris et al. 2005); vii) control of threshold conditions at the design stage by introducing adding physical devices such as tube inserts [9]; and viii) criterion of minimum sensitivity of heat exchangers to fouling effects using pinch analysis at design stage in HENs [10]. However, these mitigation strategies are not commonly applied in industry as it increases the capital and operating costs of the heat exchangers. Furthermore, the effectiveness of these strategies is not quantifiable in many cases. Regular cleanings of the heat exchangers are the only option that involves a minimum of cleaning and maintenance costs in addition to the operating costs.

Several methods for optimization of cleaning schedules for single heat exchangers have been proposed in the literature. Rodriguez and Smith [11] presented a mathematical optimization method to optimize the operating conditions to mitigate fouling in the HEN. Their study proved that the operating variables, such as wall temperature and flow velocity have significant effect on fouling deposition rate. The approach combines the optimization of operating conditions with the optimal management of cleaning actions in a comprehensive mitigation strategy and the method applied is demonstrated with a case study of a refinery crude oil preheat train. Compared with the existing strategy to mitigate fouling by managing cleaning, the proposed approach leads to higher energy savings, lower operational costs and fewer disturbances to the background process.

Panjeshahi and Tahouni [12] developed a new procedure for pressure drop optimization in debottlenecking of the HEN. This procedure enables the designer to study pump or compressor replacement whilst at the same time optimizing the additional area and operating cost of the network. It deals with the problem of optimal debottlenecking of the HEN considering the minimum operating cost. Moreover, one can consider the possibility of the replacement of a given pump with a smaller one. the new procedure have been effectively applied to a crude oil preheat train, which was subject to some 20% increase in throughput and the corresponding results proved to be accurate.

Although most of the developed mathematical models in these studies are rigorous, the HEN mathematical model can be improved by: i) Calculation of bulk properties such as density, viscosity, heat capacity and thermal conductivity based on intermediate operating conditions instead of assuming all the bulk properties are constant, ii) variation of process crude oil feed to the CPT instead of maintaining a single process crude oil feed. In conjunction with that, a more promising optimization model can be developed to determine the cleaning cycles of heat exchangers in the HEN.

1.1 Problem Statement

An optimal cleaning schedule for the HEN is required to be developed in order to improve the plant economy. However, the determination of an optimal cleaning schedule for a HEN is a challenging task due to the non-availability of reasonably accurate integrated mathematical models for the HEN under consideration and the difficulty in developing accurate fouling prediction models that follow different fouling mechanisms in different heat exchangers. A study on the development of integrated mathematical model for a HEN and the development of fouling prediction models is therefore urgently required to develop and solve a more meaningful optimisation problem to obtain an optimal cleaning schedule.

1.2 Objectives of Research

The major objectives of this research are:

a) To develop a rigorous crude preheat train simulation model

b) To develop appropriate fouling models from an industrial plant data

c) To analyze the effects of fouling on heat transfer performance of crude preheat train

d) To formulate and solve a cleaning schedule optimization problem for the crude preheat train with economic analysis

1.3 Scope of Research

The research work is divided into three major activities. Firstly, historical operating data and design data of the CPT will be collected from an operating refinery. Secondly, a rigorous integrated CPT model will be developed and calibrated to represent the performance of the plant heat exchangers. The effects of fouling on heat transfer performance of the CPT will be evaluated simultaneously. Thirdly, an optimization problem will be formulated with operational constraints (cleaning durations, duration between cleanings, maximum allowable fouling resistance, and availability of bypass facility) to determine the cleaning schedule of the CPT that results in the least operating cost.

1.4 Research Methodology

The research is started by gathering HEN design specifications and the operational data of the CPT from the plant historian of a refinery. Utilizing the data, seven units of heat exchangers after desalter with different mechanical designs, arrangements and geometries are simulated under Petro-SIM™ environment. All these heat exchangers are simulated with different preheating mediums that entered the CPT at different operating conditions and bulk properties. These standalone heat exchanger models are then integrated into a single CPT model.

Detailed calculations of tube side heat transfer coefficients, h_i, shell side heat transfer coefficients, h_o, overall heat transfer coefficients, U, and the fouling resistances, R_f, will be performed based on the established methods. The calculated data will be used in the CPT simulation model during the studies.

Two studies will be performed using the developed CPT model. In the first study, the fouling model parameters are estimated from the historical plant data. The estimated parameters will then used to determine the fouling rates of heat exchangers. By maintaining the same inlet conditions in the CPT model, the effects of fouling on the heat transfer performance of the CPT for a duration of 300 days will be studied.

In the second study, an optimization problem will be formulated to minimize the operating cost for a specified period of operation. The objective function includes the cost of energy recovered with a clean CPT, the cost of energy recovered with CPT under fouled conditions, and the cost of cleaning. Historical operating data will be used as varying operating conditions in the CPT model. The fouling rates for each heat exchanger are averaged from the historical operating data. Then, mixed integer programming approach will be used to determine the optimum cleaning schedule of the CPT for two years subject to the specified constraints. Finally, the results of optimization will be analyzed and discussed. The overall methodology of research is illustrated as in Fig. 1.1.

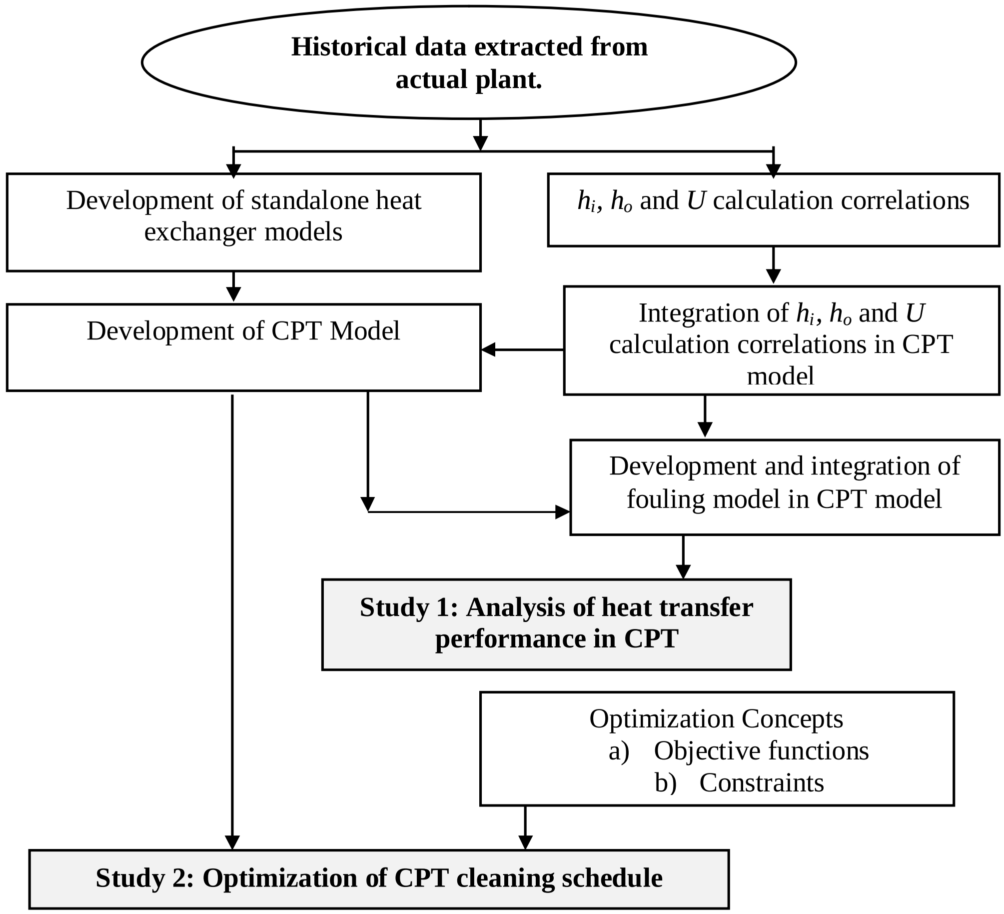

Figure 1.1: Overall Methodology of Research

1.5 Organization of the Thesis

This thesis includes five chapters. Chapter 2 presents a critical review on the literature related to the present study. In general, it covers several important aspects. These aspects are modeling, fouling predictions and cleaning schedule of HEN. Chapter 3 outlines the mathematical models used to analyze the effect of fouling in CPT. In chapter 4, the results of simulation studies for 300 days and analysis of the fouling effects on the CPT performance are presented. Lastly, chapter 5 discusses the theoretical concepts of the optimization problem to achieve cleaning schedule of CPT that have the lowest operating cost in 2 years. The optimization results are discussed and presented in this chapter.

CHAPTER 2

LITERATURE REVIEW

2.1 Introduction

Improvement in energy efficiency is very essential in process industries to reduce economic loss. A very common approach for improving energy efficiency is the integration of heat through heat exchanger networks (HENs). HENs are widely used in power plants, chemical process plants, petrochemical plants, petroleum refineries and natural gas processing plants. One of the most common examples of HEN is the crude preheat train (CPT) in refineries. It is the first processing step in refineries to preheat the crude oil before it is distilled into various petroleum products via fractional distillation.

Unfortunately, the heat exchangers in the HEN of CPT are subjected to high fouling risks and consequently the performance of these units deteriorates with time. Yeap [13] reported for a preheat train processing 100,000 bbl crude oil per day, a drop of 1 K in the coil inlet temperature (CIT) will result in approximately £23, 500 of added fuel cost each year.

The excessive economic loss due to fouling has initiated numerous studies to improve the heat recovery in HENs through various mitigation strategies. Several mitigation approaches have been applied commonly in industry. TEMA [5] recommended additional heat transfer allowance by increasing the heat transfer area during the design stage to account for fouling. This design allowance is usually a fixed value, which generally represents an asymptotic value of fouling resistance, assuming the underlying fouling process will follow an asymptotic law. However, if the fouling growth is linear with respect to time, or according to a power law or falling rate, there will be no asymptotic value. In such cases, the fouling allowance introduced at the design stage may be treated as a critical fouling resistance. The designer may have a perception that a certain time will be needed to reach this critical level of fouling and thus recommend the time between cleaning to the user. In actual operation, there is often an uncertainty concerning the extent of fouling, which should be incorporated at the design stage. It is thus important to emphasize that incorporating additional heat-transfer area does not always solve the fouling problem, but it may itself increase the problem of fouling, by introducing the changes, such as a decrease in the velocity as compared to the design value thus accelerating the fouling growth rate.

Peng et al. [6] recommended the application of continuous helical baffles instead of segmental baffles in conventional shell and tube heat exchangers for process fluids that have high fouling tendency at the shell side. The flow pattern in the shell side of the heat exchanger with continuous helical baffles was forced to be rotational and helical due to the geometry of the continuous helical baffles, which results in a significant increase in heat transfer coefficient per unit pressure drop in the heat exchanger. Properly designed continuous helical baffles can reduce fouling in the shell side and prevent the flow-induced vibration as well. The performance of the proposed heat exchangers was studied experimentally. The results indicated that the use of continuous helical baffles resulted in nearly 10% increase in heat transfer coefficient compared with that of conventional segmental baffles for the same shell-side pressure drop.

In process industries, continuous operation is vital, i.e. the heat exchangers need to be operated with highest possible availability. In these cases, a particular heat exchanger may be duplicated (i.e. a standby unit is provided), so that when one exchanger becomes excessively fouled, it can be taken out of operation for cleaning and the second exchanger can be brought into service to continuously maintain the production. This provision of a standby (or duplicate) equipment further adds to the capital cost of the plant.

The corrosion of a heat exchanger surface is also generally attributed to fouling [7]. To minimize the corrosion as well as to avoid the possibility of developing a pitting phenomenon, often more expensive materials are needed for construction, such as using titanium plates or tubes as compared to ordinary carbon steel. It is therefore expected that the cost of heat exchanger will become many times greater than that of carbon steel heat exchangers. Based on the same approach, Kukulka [8] recommends application of coating to reduce amount of fouling in heat exchangers.

Looking into design of tube bundles, Bouris et al. [14] introduced a new tube design that is called deposit determined tube shape (DDEFORM) for the development of an optimum heat exchanger arrangement that reduces gas-side fouling while increasing the heat transfer rate and decreasing the pressure drop. It was found that streamlining the shape of the tubes results in 85% lower pressure drop and 90% lower deposition rates. However, the heat transfer decreased due to reduced levels of mixing and turbulence in the modified bundles and therefore a reduction in the transverse spacing was considered in order to increase the heat transfer area and maximize the benefits of the new arrangements. The study of the new closely spaced bundles showed that they attain higher heat transfer levels with a 75% lower deposition rate and 40% lower pressure drop so that the DDEFORM tube bundle design exhibits a superior performance and presents a promising technology for the reduction of fouling in heat exchangers.

Ebert and Panchal [15] proposed the existence of threshold fouling conditions in terms of wall temperature and fluid velocity. Operation of heat exchangers below the threshold fouling conditions will increase the duration between cleanings. They proposed to control the threshold conditions at the design stage by introducing chemical additives or adding physical devices such as tube inserts. Nevertheless, these considerations might cause an increase in the capital cost due additional heat transfer area and increase in pumping cost due to increase in the pressure drop across the heat exchangers at higher fluid velocity.

Meanwhile, Brodowicz and Markowski [16] have considered the operating temperature constraint of the HEN during design stage. They proposed to design operating temperatures of heat exchangers based on criterion of a small sensitivity to fouling effects. In general, basic HEN is developed based on composite curves and minimum temperature at the pinch points. The sensitivity to fouling effects is expressed by relative capacity, β , of the heat exchanger, that is the ratio of exchanger capacity with deposits on the heat transfer, Q_f , at the end of operating period to the exchanger capacity in the absence of deposits, Q_k. The β depends on local parameters related to deposit build-up in each single heat exchanger, φ, and minimum temperature difference in the absence of deposits. Based on this approach, they concluded that the HEN is less sensitive to fouling if the values of local parameters are low. As a preventive measure, they also proposed additional compensation heat exchangers to be added to ensure all the local parameters relating to deposit build up in each heat exchanger are low.

From the angle of retrofitting the design case, Markowski [10] presented heat exchanger network retrofit using a pinch technique based approach. In this approach, the criterion of minimum sensitivity of heat exchanger to fouling effects is accounted for.

Many refineries have tried to use antifoulants to decrease the rate of fouling with some success. The use of antifoulants costs the refineries and also present problems in the downstream processing units which discourages the usage of antifoulants by the refineries. Operation of the heat exchangers below the threshold fouling conditions runs into several practical difficulties: (i) heat exchangers cannot be operated at conditions too far from the design operating conditions and (ii) operating all heat exchangers in the HEN below the fouling threshold conditions is nearly impossible. The only choice left for the refineries is to periodically clean the heat exchangers. Cleaning of each heat exchanger costs approximately USD 40,000 to 50,000 per cleaning [17]. In addition, there will be a throughput reduction when a heat exchanger is pulled out of service. Too frequent cleanings incur a huge economic loss. Inversely, less frequent cleanings reduce the energy recovery performance of the HEN. As a result, additional fuel is needed to compensate the loss of energy recovery. Therefore, an optimum cleaning schedule needs to be determined based on the cost of unrecovered energy due to fouling and the cost of cleaning and associated activities.

A cleaning schedule optimization problem for HEN requires the following components:

(i) A reasonably accurate simulation model of the HEN,

(ii) Appropriate models for predicting the fouling behavior, and

(iii) A robust optimization tool to solve the cleaning schedule optimization problem.

Only a few studies on the optimization of cleaning schedule for HENs have been reported in literature [18, 19, 20, 21,22, 23]. These studies mainly differ in the use of different types of HEN simulation model, fouling prediction models and the optimization techniques.

This chapter will present the relevant literatures on the HEN simulation models, fouling models and the formulation of cleaning schedule optimization problem together with the solution techniques.

2.2 Modeling of Heat Exchanger Network

Mathematical model is essential to accurately assess the performance of HEN by relating the total heat transfer rate to quantities such as the inlet and outlet fluid temperatures, the overall heat transfer coefficient, and the total surface area for heat transfer.

2.2.1 Single Heat Exchanger

A model of single heat exchanger consists of energy conservation and heat-transfer equations [24]. The energy conservation equation with assumption of negligible heat loss to the surroundings and no phase change is given by

$$Q = M_h C_{p,h} \left(T_{h,i} - T_{h,o} \right) = M_c C_{p,c} \left(T_{c,o} - T_{c,i} \right) \tag{2.1}$$

where

$C_{p,h}$ = heat capacity rates of the hot fluid

$C_{p,c}$ = heat capacity rates of the cold fluid

M_h = mass flow rate of the hot fluid

M_c = mass flow rate of the cold fluid

T_h = hot fluid terminal temperatures (inlet, *i* and outlet, *o*)

T_c = cold fluid terminal temperatures (inlet, *i* and outlet, *o*)

The average driving force for heat transfer is the log-mean temperature difference, *LMTD*, when two fluid streams are in counter-current flow. It depends on the flow arrangement and degree of fluid mixing within each fluid stream for the heat exchanger. It is given as

$$LMTD = \frac{(T_{h,o} - T_{c,o}) - (T_{h,i} - T_{c,i})}{\ln \frac{(T_{h,o} - T_{c,o})}{(T_{h,i} - T_{c,i})}} \tag{2.2}$$

However, the overriding importance of other design factors causes most heat exchangers to be designed in flow patterns different from true counter-current flow. The true *LMTD* of such flow arrangements will differ from the *LMTD* by a correction factor, *F*, that depends on the flow pattern and terminal temperatures. The heat transfer equation incorporating *F* is given as

$$Q = UAF.LMTD \tag{2.3}$$

where

A = effective heat transfer area

F = correction factor for *LMTD*

U = overall heat transfer coefficient

The overall heat transfer coefficient, *U*, is related to the fouling resistances for both the shell and tube sides of the heat exchangers, R_i and R_o, and is given as

$$\frac{1}{U}=\frac{d_o}{d_i h_i}+\frac{d_o R_i}{d_i}+\frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2k_w}+R_o+\frac{1}{h_o} \tag{2.4}$$

where

d_i = inner diameter of the tube

d_o = outer diameter of the tube

h_i = heat transfer coefficients of tube side

h_o = heat transfer coefficients of shell side

k_w = thermal conductivity of the tube wall material

2.2.2 Heat Exchanger Network

Heat exchanger networks are generally formed by the combination heat exchangers in series and/or parallel arrangement. Therefore, the mathematical models of the HENs are developed in a modular form. Fouling of a heat exchanger in the HEN reduces the heat transfer performance resulting in different outlet temperatures of the process streams. Consequently, the other heat exchangers downstream of HEN with the same process streams will also be affected. Hence, computations of these inlet temperatures are essential to analyse the operation of heat exchangers and identify the effects of fouling on its heat transfer rates. To compute these inlet temperatures, each heat exchanger model is developed by the set of Eqs. 2.1-2.4.

The outlet temperatures of hot and cold streams of each heat exchanger in the HEN can be estimated by solving Eqs. 2.1-2.4. An analytical solution for the outlet temperatures is obtained as

$$T_{c,o}=\left[\frac{k_1(\exp(-k_2F(k_1-1))-1}{\exp(-k_2F(k_1-1))-k_1}\right]T_{h,i}+\left[\frac{(1-k_1)(\exp(-k_2F(k_1-1))}{\exp(-k_2F(k_1-1))-k_1}\right]T_{c,i} \tag{2.5}$$

$$T_{h,o}=\left[\frac{(\exp(-k_2F(k_1-1))-1}{\exp(-k_2F(k_1-1))-k_1}\right]T_{c,i}-\left[\frac{k_1-1}{\exp(-k_2F(k_1-1))-k_1}\right]T_{h,i} \tag{2.6}$$

where

$$k_1 = \frac{M_h C_{p,h}}{M_c C_{c,h}}$$

$$k_2 = \frac{UA}{M_h C_{p,h}}$$

The flow rates of the fluid streams, M_h and $M_{c,}$ are assumed to be constant for heat exchangers that are in series and equally divided if the heat exchangers are arranged in parallel. Essential physical properties such as thermal conductivities, viscosities, densities and heat capacities are estimated based on the intermediate temperatures. In almost all the studies that were reported recently, these physical properties at intermediate streams of heat exchangers are assumed constant regardless the occurrence of fouling [18, 19, 20, 22]. This assumption resulted in constant heat transfer coefficients for both the tube and shell sides of heat exchanger in spite of the fouling. Hence, it is essential to compute the accurate physical properties for the fluids that enter the intermediate heat exchangers in the CPT.

The fouling resistance of the heat exchangers were either obtained from constant fouling rates of historical data or estimated from the fouling models. With these approaches, the effects of fouling on each heat exchanger in HEN were analysed.

For heat exchangers that are arranged in series without recycle stream, sequential solution approach is used to determine the thermal effectiveness of each heat exchanger. The thermal effectiveness of the first heat exchanger is solved followed by the subsequent heat exchangers. Simultaneous solution approach is required for HEN in which the shells are arranged in series with recycle streams, which is commonly seen in CPT of refineries. In CPT, normally hot residue from crude distillation column (CDU) is first used to preheat crude oil in last heat exchanger, followed by the second heat exchanger at downstream of desalter.

2.3 Effects of Fouling on Heat Exchanger Network

Deriving from the energy conservation and heat-transfer equations, different techniques are used by researchers to analyse the effect of fouling on HEN performance. Markowski and Urbaniec [21] developed a graphical approach to investigate the effects of deposits build up in HEN, and its interaction with antecedent exchangers. Based on the graphical approach, outlets temperatures of heat exchangers without deposits are compared against the outlets temperatures of heat exchanger with the deposit build up. The authors analysed the effect of fouling in the HEN comprising of 10 heat exchangers arranged in series using the graphical approach.

The formal introduction of the ε *-NTU* method for the heat exchanger analysis was by London and Seban [25] to evaluate the thermal effectiveness of HEN. In this method, the total heat transfer rate from the hot fluid to cold fluid in the exchanger is expressed as

$$Q = \varepsilon C_{\min}(T_{h,i} - T_{c,i}) \tag{2.7}$$

where

ε = effectiveness of a heat exchanger

$C_{\min}$ = lower thermal capacity rate of the hot or cold streams

In a two-fluid heat exchanger, one of the streams will usually undergo a greater temperature change than the other. The first stream is said to be the “weak” stream, having the lower thermal capacity rate, $C_{\min}$, and the other with higher thermal capacity rate, $C_{\max}$, which is the “strong” stream. Number of transfer units, *NTU*, is proposed to represent the thermal effectiveness of each heat exchanger in the HEN in the presence of fouling. It is defined as the ratio of the overall conductance to the smaller heat capacity rate

$$NTU = \frac{1}{C_{\min}}\left[\frac{1}{\frac{1}{(\eta hA)_h} + R_1 + R_w + R_2 + \frac{1}{(\eta hA)_c}}\right] \tag{2.8}$$

where

R_1 = thermal resistances due to fouling on the hot side

R_2 = thermal resistances due to fouling on the cold side

η = heat transfer efficiency

h = heat transfer coefficient

A = area of heat transfer

ψ-P method was originally proposed by Smith [26] and modified by Mueller [27] to evaluate the heat exchanger thermal effectiveness of HEN. In this method, a new term ψ is expressed as the ratio of the true mean temperature difference to the inlet temperature difference of the two fluids.

$$\psi = \frac{LMTD}{T_{h,i} - T_{c,i}} \tag{2.9}$$

The *LMTD* method can also be used when the inlet and outlet temperatures of both the hot and cold fluids are known. When the outlet temperatures are not known, the *LMTD* can only be used in an iterative scheme. In this case the effectiveness-*NTU* method can be used to simplify the analysis. Arsenyeva [28] determined the optimal design for multi-pass plate and frame heat exchanger with mixed grouping of plates using the ε-*NTU* method. The heat transfer effectiveness is used as objective function with inequality constraints. The optimizing variables include the number of passes for both streams, the number of plates with different corrugation geometries in each pass, the plate type and size. Similarly, Sanaye and Niroomand [22], used ε-*NTU* method to determine the initial outlet temperatures of intermediate hot and cold streams for every exchanger in the HEN that are arranged in series.

Jeronimo et al. [29] have utilized *P-NTU* method to evaluate the performances of heat exchangers in an oil refinery, based on the assessment of the *NTU* and thermal efficiencies. The effects of changing the feedstock charge, particularly the flow rates of the fluids, are taken into account. The data required for the *P-NTU* calculation are the four inlet/outlet temperatures of the heat exchanger unit and one of the flow rates. *P-NTU* represents a variant of the ε-*NTU* method. The temperature effectiveness of the tube side fluid for a shell and tube heat exchanger is referred as “thermal effectiveness” of the tube side fluid regardless it is hot or cold fluid. It is defined as the ratio of the temperature rise or drop of the tube side fluid to the difference of the inlet temperature of the two fluids. According to this definition, P is given by

$$P = \frac{t_2 - t_1}{T_1 - t_1} \tag{2.10}$$

where

t_1 = tube side inlet temperature

t_2 = tube side outlet temperature

T_1 = shell inlet temperature

At low *NTU*, the exchanger effectiveness is generally low. With increasing values of *NTU*, the exchanger effectiveness increases, and in the limit it approaches asymptotic value. However, there are exceptions where the effectiveness decreases with increasing *NTU* after reaching a maximum value.

All the studies show that the overall heat transfer coefficient, *U*, decreases with increase of fouling for single heat exchanger. Nevertheless, this rule cannot be generally applied to heat exchangers in the HEN with different physical arrangements. The heat transfer performance deteriorates in some of the heat exchangers while marginal increase is observed in other heat exchangers. None of the research focus on investigation of the marginal increase of heat transfer performance in some of the heat exchangers in HEN under fouled conditions.

2.4 Fouling Prediction Models

Accurate fouling models are essential to predict the heat transfer performances for the determination of optimal cleaning schedule of the HEN. Development of fouling models to explain the complex nature of fouling process is difficult due to the multiple fouling mechanisms associated with crude oils. Numerous research works have attempted to develop mathematical models ranging from theoretical, empirical, to semi-empirical models.

2.4.1 Theoretical Models

Theoretical fouling models were developed to carry out systematic investigation on fouling. Ctittenden and Kolaczkowski [30] proposed a general model that considers the transport of fouling precursors and chemical reaction. The proposed modified model was able to determine the polystyrene deposition from the dilute solution of styrene in kerosene. The equation is given by

$$\frac{dR_f}{dt} = \frac{1}{\rho_f k_f}\left[\frac{C_{rb}}{\frac{\rho(d_i - 2x)^{1.8}(Sc_r)^{0.67}}{1.213k\mu^{0.2}M^{0.8}} + \frac{1}{Ae^{\frac{-E}{RT}}}} - \frac{1.213k\mu^{0.2}M^{0.8}C_{Di}}{\rho(d_i - 2x)^{1.8})(Sc_D)^{0.67}}\right] \tag{2.11}$$

where

C_{rb} = concentration of precursors

M = mass flow rate

Sc = Schmidt number (reactant, *r* and deposit, *D*)

C_{Di} = foulant concentration at the solid-liquid interface

x = thickness of deposit

Srinivasan and Watkinson [31] developed a fouling correlation using a modified film temperature, $T_f^{'}$, weighted more heavily on the surface temperature and is given by

$$\frac{dR_f}{dt} = av^{-0.35}\exp(\frac{-E}{RT_f^{'}}) \tag{2.12}$$

$$T_f^{'} = 0.3T_b + 0.7T_S$$

where

a = constant

T_b = bulk temperature

T_S = surface temperature

The Eq. 2.12 was found able to fit the fouling data within $\pm 8\%$ accuracy based on the fouling experiment using Canadian crude oils in a recirculation fouling loop that equipped with an annular electrically heated probe.

Although the theoretical models are based on fundamental principles, they cannot be used successfully to describe the fouling process due to the requirements of many parameters that are difficult to determine especially in operational refineries. These parameters include concentration of precursors, Schmidt number, foulant concentration at the solid-liquid interface and thickness of deposits on tube surface. Due to these limitations, researchers are working on either the empirical models or semi-empirical models that have less dependency on the parameters as above [17].

2.4.2 Empirical Models

Kern and Seaton [32] observed from the industrial data that the fouling resistance of a heat exchanger increases asymptotically with the assumptions that the heat exchangers process the same fluids type and inlet conditions (temperature, pressure and flow) are held constant. The fouling resistance, R_f, increases following a first order response and depends on the time constant of the fouling formation, τ, and asymptotic fouling resistance, R_f^∞. The values of R_f^∞, and τ, must be determined from the historical data. The asymptotic fouling model is

$$R_f(t) = R_f^\infty (1 - \exp(-t/\tau)) \tag{2.13}$$

where

t = time since the last cleaning

Sanaye and Niroomand [22] applied the asymptotic fouling model to estimate the fouling behavior of heat exchangers in HEN. The fouling parameters, R_f^∞ and τ were estimated based on empirical data obtained from the operating inlet and outlet temperature variations of the hot and cold streams recorded for the various heat exchangers in the Khorasan Petrochemical Plant for 2 years. The results show good consistency between the exchanger outlet temperature predicted by the model and real operating conditions of the system. The model does not consider the effects of important parameters such as fluid type, flow temperature and velocity of the fluids.

Besides asymptotic fouling model, computational intelligence techniques such as artificial neural networks (ANN) are used to simulate HEN models that have complex relationship between inputs and outputs, and the patterns of collected data are hard to determine based on semi-empirical calculations. Aminian and Shahhosseini [33] applied the ANN modeling based on a refinery design and operating data. The trained ANN model predicted the fouling rates with an overall mean relative error (OMRE) of 26.23% which was much better than the predictions by the semi-empirical models. Radhakrishnan et al. [34] also applied ANN based fouling prediction model for CPT of CDU and presented a preventive maintenance scheduling tool using historical operating data.

2.4.3 Semi Empirical Models

Kern and Seaton [32] proposed a simple concept to model fouling processes that involved chemical reactions. They assumed that the rate of fouling is defined as difference between the rates of deposition, m_d, and rate of removal, m_r with the assumptions: i) no chemical reaction is taking place, ii) net deposition is the result of deposition minus fouling removal, iii) fouling removal increases with mass of deposit and iv) the rate of deposition is independent of mass of deposit. The correlation is given by

$$\frac{dR_f}{dt} = m_d - m_r \tag{2.14}$$

Ebert and Panchal [15] is the first to develop the threshold fouling model for the HEN as a semi-empirical basis to describe the rate of fouling formation in heat exchangers. Their fouling model assumed that: i) the net deposition is given by formation minus removal of foulant from the thermal boundary layer, ii) foulant is formed in the boundary layer by reactions which can be grouped as one step reaction, iii) concentration gradients of reactants in the boundary layer is negligible, iv) foulant is transported by diffusion and turbulent eddies from the boundary layer to the bulk flow, v) temperature profile in the boundary layer is linear and, vi) an integrated reaction term can be expressed by the film temperature in the boundary layer.

The fouling model is controlled by the deposition mechanism and the suppression mechanism to predict the rate of fouling. The first mechanism is related to the formation of deposits due to chemical reactions. The second mechanism considers the shear stress at the tube surface, which acts to mitigate fouling. The expression for predicting the linear fouling rate is given by

$$\frac{dR_f}{dt} = \alpha \operatorname{Re}^{\beta} \exp\left(\frac{-E}{RT_f}\right) - \gamma \tau_w \tag{2.15}$$

Panchal et al. [35] subsequently modified Eq. 2.16 to cover a wider range of physical parameters by considering $Pr^{-0.33}$ parameter in the deposition term. The modified expression is

$$\frac{dR_f}{dt} = \alpha \operatorname{Re}^{\beta} \operatorname{Pr}^{-0.33} \exp\left(\frac{-E}{RT_f}\right) - \gamma \tau_w \tag{2.16}$$

Polley et al. [36] later proposed an alternative form of the threshold model. Modifying the basic equations as proposed by Panchal et al. [35], the removal term is linked to the rate of convective mass transfer from the surface to the bulk liquid. The detail modifications are: (i) The film temperature term is replaced with the wall temperature in the Arrhenius term; (ii) The removal term is assumed proportional to $Re^{0.8}$. The model is given by

$$\frac{dR_f}{dt} = \alpha' \operatorname{Re}^{-0.8} \operatorname{Pr}^{-0.33} \exp\left(\frac{-E}{RT_w}\right) - \gamma' \operatorname{Re}^{0.8} \tag{2.17}$$

2.5 Cleaning Schedule Optimisation

Cleaning schedule optimizations of heat exchanger or HEN have been carried out in industry. The optimization problem is formulated to determine the best cleaning timing where a balance has to be reached between; i) additional fuel cost to compensate the decreasing heat transfer performance: ii) additional pumping cost to overcome high pressure drop across the heat exchangers due to presence of deposits that restricts the flow of fluids: iii) cleaning and maintenance costs of heat exchanger due to fouling, iv) production loss due to insufficient heat transfer to maintain the desired temperature of process fluid, and v) production loss due to reduction of plant throughput.

The cleaning schedule for single heat exchangers has been proposed in the literatures. Epstein [37] worked on an analytical method to estimate the optimal evaporator cycle with scale formation. Casado [38] used a cost model to calculate the optimal cleaning cycle of a heat exchanger under fouling by exploring the major operating trade off. Based on these works, Sheikh et al. [39] presented a reliability based cleaning strategy by incorporating uncertainty in a linear fouling model. These optimization problems merely considered linear relations for the impacts of the different costs to the heat exchanger due to fouling.

In process plants, complex HENs exist, in which the interactions between the heat exchangers within HEN due to fouling is complex. Kotjabasakis and Linhoff [40] presented a network sensitivity analysis technique to gauge the effect of fouling on networks. However, this technique cannot readily extend to the case of cleaning, where the network configuration is temporarily changed due to cleaning of one or more heat exchangers. The interactions of operating parameters and the effects of fouling between heat exchangers cannot be considered as well.

Thus, a solution of the scheduling problem always involve the requirements of: 1) a HEN model that is able to determine the interactions of operating parameters between the heat exchangers, 2) a fouling model to estimate the fouling rates, and 3) an optimization problem formulation to achieve the optimum cleaning schedule. The objective function of the optimization problem can be formulated in several forms, such as the maximization of outlet temperature of the last heat exchanger in HEN, maximization of heat transferred, maximization of production rate, minimization of costs or combinations of any of the forms. Equipped with an accurate HEN and fouling models, numerical techniques such as integer linear programming (MILP) or mixed integer non-linear programming (MINLP) can be used to solve the optimization problem while satisfying the operational and maintenance constraints. These constraints are important for the robust and accurate convergence of the formulated optimization problem.

Smaili et al. [18] formulated an optimization problem to obtain cleaning schedule for the thin juice preheat train in a sugar refinery. Preheat train case study consisted of three heat exchangers arranged in series, followed by four heat exchangers arranged in parallel, and lastly four heat exchangers downstream that are arranged in series. The optimization problem was formulated under the assumptions; i) constant inlet temperature and mass flow rates, ii) external control options prompted by other sections of the plants were not incorporated, iii) constant linear fouling rates obtained from 150 days of reconciled data. The objective function is set to minimize the consumption of selected preheating utility. The constraints applied for this optimization problem are: i) binary variables that describe the status of the heat exchanger, ii) the outlet temperature of last heat exchanger must satisfy minimum process specification, and iii) only one heat exchanger is cleaned at a time. This approach is reported very efficient for the solution of overall MINLP problem under GAMS modeling.

Motivated by the above practical industrial problem, the simultaneous consideration of scheduling and process operation aspects seem to be promising. Georgiadis and Papageorgiou [19] studied the optimum energy recovery and cleaning management in HEN of milk processing under fouling. MILP algorithm was used to solve the objective function that considered only additional utility and cleaning costs. The former term is assumed to be a linear function of utility temperature while the later term considered the cleaning cost of heat exchangers when it is offline. Similarly, Smaili et al. [18] used binary variables in the objective function to indicate the operating status of the heat exchangers.

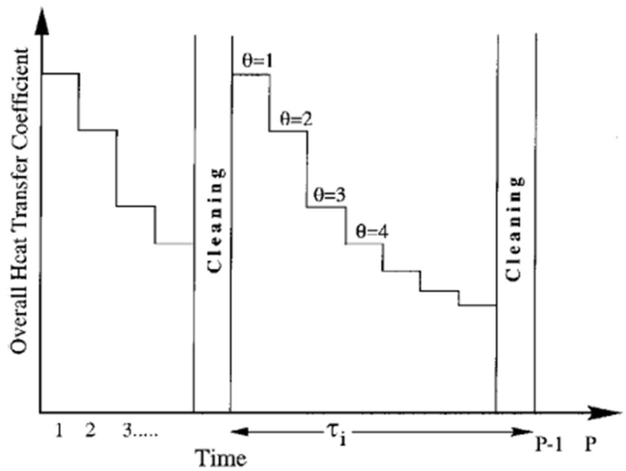

Figure 2.1: Non-linear decaying profile of overall heat transfer coefficient over 1 cycle [19].

The assumptions to ensure the convergence of optimization problem are made, viz: i) all the operating conditions to the inlets of the HEN model are constant, ii) all the physical properties in between the heat exchangers in HEN were assumed constant: iii) Non-linear decaying profile of overall heat transfer coefficients are estimated from historical data for the heat exchangers over the period of optimization, iv) the optimal utility utilization profile over time, and v) the maximum number of cleaning operations required is fixed. The formulated optimization problem is solved for serial HENs, parallel HENs, combination of serial and parallel HENs and network arrangements arising from the combination of these basic operations. It is proven in this optimization study that availability of hot utility in the plant plays an important role to minimize the operating cost of HEN in milk processing. The formulations presented in this paper are suitable for serial and parallel HENs, as well as network arrangements arising from the combinations of these two arrangements. Nevertheless, the proposed models cannot directly be applied to networks with stream splits. On the other hand, the developed MILP algorithm has very tight constraints and cannot guarantee global solutions especially with solution of problem with more than 10 heat exchangers.

Markowski and Urbaniec [21] formulated cleaning schedule optimization problem for HEN that consisted of 10 heat exchangers arranged in series for a continuous plant operation. Graphical approach is adopted to illustrate the temperature distribution in the HEN and impacts of deposit build up in the antecedent exchangers. The assumptions made for this optimization problem are: i) all the inlet operating parameters and flows to the HEN model were constant, ii) Constant linear fouling rates from literature are used for all the heat exchangers, iii) Duration of production period is one year. The online cleaning is assumed for all the heat exchangers in HEN considered the interventions of cleaning schedules and diagnostic of deposit build up of the heat exchangers in HEN. The objective function of the formulated problem is to maximize the heat transfer performance while minimizing the cleaning cost of the HEN satisfying the time constraint imposed on time intervals between cleaning interventions. The heat transfer performance was evaluated based on the difference of energy recovered depending on the operating status of the heat exchangers. It was observed that solving this MINLP problem required large computational effort as there are both integer and continuous decision variables. Besides, this approach is only applied for HEN with heat exchangers arranged in series. In the example of the HEN comprising of 10 heat exchangers, the cleaning interventions scheduled is able to save 5% of the maximum attainable value of energy recovered in the HEN.

Slightly different methodology was adopted by Sanaye and Niroomand [22] to determine optimum cleaning schedule of HEN in a petrochemical plant. The HEN mathematical model consisted of a set of nonlinear equations describing the performance of each heat exchanger and a set of linear equations describing the links of the units and the splitting/mixing of the streams. The asymptotic fouling model, with its parameters extracted from a petrochemical plant for a period of two years is used to estimate the fouling behavior of heat exchangers. Integrating the mathematical and asymptotic fouling models, an objective function is formulated to minimize network operating cost during a specific duration. These operating costs include i) unrecovered energy cost that was the difference between the heat transfer performance of the heat exchanger when it is under clean and fouled conditions, and ii) the cleaning and maintenance costs incurred when the heat exchanger was offline for cleaning. Bypassing and cleaning of all heat exchangers was studied to investigate the effect of cleaning to each individual heat exchanger and the HEN. In this optimization problem, it is assumed that: i) all the inlet operating parameters and flows to the HEN model were constant, ii) physical properties of the intermediate process streams are constant, and iii) minimum successive cleaning interval for each heat exchanger. Based on an optimization period of two years, savings were achieved and reported.

Rodriguez and Smith [11] have studied on the optimization of HEN cleaning schedule in a different approach. Instead of focusing full attention on heat transfer performance and operational costs, the study also emphasized on changes in operating conditions of existing networks consisted of eleven heat exchangers arranged in series to reduce the severity of fouling. A simplified CPT model without split streams was used as case study. The *P-NTU* method was used to evaluate the heat transfer performance of the heat exchangers in HEN depending on the ratio of heat capacity flow rates, the number of transfer units and the thermal effectiveness factor. Fouling model as proposed by Polley et al. [36] was used to estimate the fouling deposition, which is considered to be merely happening at crude oil side of the heat exchangers. It is assumed that every heat exchanger can be taken out for cleaning service. With these models and assumptions, the optimization problem was formulated to find the optimal combination of cleaning actions and operating conditions that minimizes the total operating cost of the HEN, which comprises the cleaning and cumulative energy costs. The cumulative energy cost corresponded to the additional energy consumption and energy penalties due to fouling. The objective function comprises of continuous variables, representing the setting of the manipulated operating variables and binary variables, representing the cleaning schedule. The optimization problem was solved using simulated annealing (SA) optimization method and was able to demonstrate that the proposed approach leads to higher savings as compared to optimization results that solely based on cost factors. However, the changes in the operating conditions requires the reduction of heat transfer performance in most of the cases, which led to increase of utilities consumption to maintain the outlet temperature of HEN at their targeted value.

2.6 Summary

Among the fouling models, Panchal et al. [35] model is the most commonly used represent fouling in HENs. Despite numerous modeling approaches have been adopted for the mathematical modeling of HEN, none of the studies were carried out using rigorous mathematical model under a simulation platform, in which the inputs to the model are based on dynamic operating conditions that are extracted from an actual operating process plant. Only limited studies have been reported to quantify the effects of fouling on each heat exchanger in the HEN.

Numerous methodologies have been studied to optimize the cleaning schedule of HEN for either online cleaning or offline cleaning. As reported, it is crucial for the operation to predict the performance of the heat exchangers in the HEN, and schedule the cleaning cycles of heat exchangers to minimize the total operating cost. This is essential to production planning of the refinery in longer run to avoid emergency shutdown of the plant due to sudden drop of HEN performance.

CHAPTER 3

CRUDE PREHEAT TRAIN SIMULATION MODEL

3.1 Introduction

One of the major fouling mitigation approaches involves cleaning of the heat exchangers based on an optimum cleaning schedule. In order to develop such an optimum cleaning schedule, the development of an accurate and a rigorous simulation model of the CPT is essential. A CPT consists of many heat exchangers arranged in parallel and series configurations and each heat exchanger is described by the same mathematical model with different values of parameters and operating conditions. Development of a mathematical model for a single heat exchanger will be explained in detail which are then duplicated and connected to represent the industrial CPT configuration.

In this chapter, the development of a simulation model for a crude preheat train post desalter in an operating refinery will be discussed in detail. The mathematical model of shell and tube heat exchanger is presented in Section 3.2. The design details and design operating conditions of the heat exchangers in the CPT and the building up of the CPT simulation model under Petro-SIM™ environment are provided in Section 3.3. Development of semi-empirical fouling models for each heat exchanger in the CPT using the operational data from the plant is discussed in Section 3.4. Section 3.5 presents the integration of the CPT simulation model with the fouling models.

3.2 Modeling

In the CPT considered in this study, shell and tube heat exchangers with multiple shell- and tube-side passes are used. The underlying equations that describe the heat transfer in these heat exchangers are explained in the following section.

3.2.1 Governing Equations

Consider an infinitesimally small volume element in a counter-current shell and tube heat exchanger as shown in Fig. 3.1. The hot fluid flows in shell side with a mass flow rate of M_h. The hot fluid enters the volume element at a temperature of T_h^i and leaves at a temperature of T_h^o. The cold fluid flows through the tube side with a mass flow rate of M_c. The cold fluid enters and leaves the volume element at temperatures of T_c^i and T_c^o, respectively.

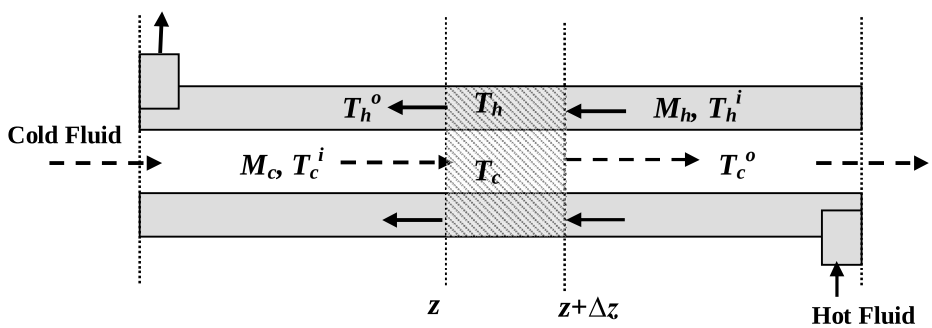

Figure 3.1 Schematic diagram of heat exchanger in a counter-flow arrangement.

Regardless of the flow configuration and geometry, an energy balance can be written for the volume element. It is assumed that: i) the differences in the kinetic and potential energies of the fluid streams between the inlet and outlet of the volume element are negligible, ii) the heat exchanger is assumed to be at steady state, with temperatures of both hot and cold fluids at equilibrium and equal to the temperatures of fluids that exit the volume element in the shell and tube sides, iii) the heat loss to surroundings through heat exchanger is negligible, and iv) the fluids are always in liquid phase in both the tube and shell sides. With these assumptions, the energy balance for both the tube and shell sides can be written as

$$M_cC_{p,c}(T_c^o - T_{ref}) = M_cC_{p,c}(T_c^i - T_{ref}) + Q \tag{3.1}$$

$$M_hC_{p,h}(T_h^o - T_{ref}) = M_hC_{p,h}(T_h^i - T_{ref}) - Q \tag{3.2}$$

where

$C_{p,c}$ = heat capacity of the cold fluid

$C_{p,h}$ = heat capacity of the hot fluid

Q = heat transfer rate

The rate of heat transfer, Q, across the heat transfer surface of the control volume is given by

$$Q = UA_o(T_h^o - T_c^o) \tag{3.3}$$

where

U = the overall heat transfer coefficient

The heat transfer area, A_o, is given as

$$A_o = \pi d_o \Delta z \tag{3.4}$$

where

Δz = length of small element region

d_o = outer diameter of the tube

Combining the Eqs. 3.1, 3.2 and 3.3, results in Eq. 3.5 for tube side and Eq. 3.6 for shell side

$$\frac{T_c^o - T_c^i}{\Delta z} = \frac{\Delta T_c}{\Delta z} = \frac{UA_o(T_c^o - T_h^o)}{M_cC_{p,c}} \tag{3.5}$$

$$\frac{T_h^i - T_h^o}{\Delta z} = \frac{\Delta T_h}{\Delta z} = \frac{UA_o(T_h^o - T_c^o)}{M_hC_{p,h}} \tag{3.6}$$

For the limiting case of $\Delta z \to 0$, the difference equations of Eqs. 3.5 and 3.6 result in

$$\frac{dT_c}{dz} = \frac{UA_o(T_c^o - T_h^o)}{M_c C_{p,c}} \tag{3.7}$$

$$\frac{dT_h}{dz} = \frac{UA_o(T_h^o - T_c^o)}{M_h C_{p,h}} \tag{3.8}$$

3.2.2 Overall Heat Transfer Coefficient

The overall heat transfer coefficient, U, is given by

$$\frac{1}{U} = \frac{d_o}{d_i h_i} + \frac{d_o R_i}{d_i} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2k_w} + R_o + \frac{1}{h_o} \tag{3.9}$$

where

d_i = inner diameter of the tube

R_i = fouling resistances at the tube side

R_o = fouling resistances at the shell side

k_w = thermal conductivity of the tube wall material

3.2.2.1 Tube Side Heat Transfer Coefficient

The tube side heat-transfer coefficient, h_i, for turbulent flow conditions are estimated using Colburn method [41] as

$$h_i = 0.023 \frac{\mathrm{Re}^{0.8}\,\mathrm{Pr}^{0.4}\,k}{d_i} \qquad 2100 < \mathrm{Re} < 10000 \tag{3.10}$$

where

Re = Reynolds number of the fluid

Pr = Prandtl number of the fluid

k = thermal conductivity of the fluid

3.2.2.2 Shell Side Heat Transfer Coefficient

Bell-Delaware method [42] is used to calculate the shell side heat transfer coefficient, h_o, for heat exchangers with vertical segmental baffles. The h_o is a function of ideal heat transfer coefficient, h_{ideal}, and correction factors as

$$h_o = h_{ideal} J_c J_b J_l J_s J_r \tag{3.11}$$

where

J_c = correction factor for baffle cut and spacing

J_l = correction factor for baffle leakage effects, including both shell-to-baffle and tube-to-baffle leakage

J_b = correction for the bundle bypass flow

J_s = correction for variable baffle spacing in the inlet and outlet sections

J_r = correction for adverse temperature gradient buildup in laminar flow

The ideal heat transfer coefficient, h_{ideal}, is given as

$$h_{ideal} = J_i C_{p,s} \left(\frac{C_p . \mu}{k} \right)^{-2/3} \left(\frac{M_s}{S_m} \right) \left(\frac{\mu_s}{\mu_w} \right)^{0.14} \tag{3.12}$$

$$J_i = 0.236 \left(\frac{\rho v d_o}{\mu} \right)^{-0.346} \qquad Re \geq 1000$$

where

M_s = fluid mass flow at shell side

S_m = overall shell side cross flow area

For the baffle cut range 15-45 percent, segmental baffle window correction factor, J_c, is expressed by

$$J_c = 0.55 + 0.72 F_c \tag{3.13}$$

where

F_c = pure cross flow area of the shell

The correction factor of baffle leakage effects for heat transfer, J_l, is given by

$$J_l = 0.44(1-\frac{S_{sb}}{S_{sb}+S_{tb}}) + \left(1-0.44\left(1-\frac{S_{sb}}{S_{sb}+S_{tb}}\right)\right)e^{-2.2\frac{S_{sb}+S_{tb}}{S_m}} \tag{3.14}$$

where

S_{sb} = total tube-to-baffle leakage area of shell

S_{tb} = shell-to-baffle leakage area within the circle segment occupied by the baffle

S_m = overall cross flow area of the shell

Correction factor of bundle bypass effects for heat transfer, J_b, is determined by the number of sealing strips (pairs) in one baffle, N_{ss}, and the number of tube rows that crossed between baffle tips in one baffle section, N_{tcc}. The expression for J_b is correlated as

$$J_b = \exp\left(-C_{bh}\frac{S_b}{S_m}\left(1-\left(2\frac{S_{sb}}{S_{sb}+S_{tb}}\right)^{1/3}\right)\right) \tag{3.15}$$

C_{bh} = 1.25 for laminar flow, $Re \leq 100$

= 1.35 for turbulent and transition flow, $Re > 100$

where

S_b = ratio of the bypass area

The correction factor for variable baffle spacing in the inlet and outlet sections, J_s, is given as

$$J_s = \frac{(N_b - 1) + \left(\frac{L_{bi}}{L_{bc}}\right)^{1-n} + \left(\frac{L_{bo}}{L_{bc}}\right)^{1-n}}{(N_b - 1) + \left(\frac{L_{bi}}{L_{bc}} - 1\right)^{1-n} + \left(\frac{L_{bo}}{L_{bc}} - 1\right)^{1-n}} \tag{3.16}$$

n = 0.6 for turbulent flow

where

N_b = number of baffles

L_{bi} = central baffle spacing

L_{bc} = exact central spacing

L_{bc} = outlet baffle spacing

Heat transfer correction factor for adverse temperature gradient in laminar flow, J_r, applies only if the shell-side *Re* is less than 100 and is fully effective only if deep laminar flow characterized by *Re* is less than 20. J_r can be expressed as

If $Re < 20$

$$J_r = \frac{0.55 + 0.72F_c}{\left(\left(N_{tcc} + N_{tcw}\right)\left(N_b + 1\right)\right)^{0.18}} \tag{3.17}$$

If $20 < Re < 100$

$$J_r = \frac{1.51}{\left(\left(N_{tcc} + N_{tcw}\right)\left(N_b + 1\right)\right)^{0.18}} + \left(\frac{20 - \mathrm{Re}}{80}\right)\left(\frac{1.51}{\left(\left(N_{tcc} + N_{tcw}\right)\left(N_b + 1\right)\right)^{0.18}} - 1\right)$$

(3.16)

If *Re* = 100

$$J_r = 0.4 \tag{3.18}$$

If *Re* >100

$$J_r = 1.0 \tag{3.19}$$

Shell side heat transfer coefficient, h_o, for continuous helical exchanger is computed using the Peng method [6]. The equation is expressed as

$$\frac{h_o d_o}{k} = 0.0451\left(\frac{\rho v d_i}{\mu}\right)^{0.7}\left(\frac{C_p \mu}{k}\right)^{1/3} \tag{3.20}$$

3.2.2.3 Fouling Resistance

The fouling layers on the inside and the outside surfaces are known to increase with time as the heat exchanger is operated. The total fouling resistance, R_f, to heat transfer between hot and cold streams is the difference between the inverses of actual and clean values of overall heat transfer coefficients. The R_f is expressed as

$$R_f = \frac{1}{U_f} - \frac{1}{U_c} \tag{3.21}$$

3.2.3 Physical Properties

Physical properties of cold and hot fluids are essential to determine heat transfer duty of heat exchangers in the HEN. In single unit of heat exchanger, in which inlet temperatures and one of the outlet temperatures of hot or cold fluids are known, physical properties can be estimated based on the average bulk temperature. There are difficulties in applying this approach to HEN. In HEN, except the first heat exchanger, both the inlet and outlet temperatures of intermediate heat exchangers vary depending on the calculated outlet temperatures from upstream heat exchangers. Thus, accurate calculation of heat transfer performance cannot be performed if constant values of physical properties are used for these intermediate heat exchangers due to the changes in inlet and outlet temperatures caused by the fouling. Accurate prediction of physical properties for the fluids is essential in this study. Due to this, it is important to always estimate the physical properties of the intermediate heat exchangers using the first principal model.

Calculation of three PVT derivatives ($\left(\frac{\delta P}{\delta V}\right)_T, \left(\frac{\delta T}{\delta P}\right)_V, \left(\frac{\delta V}{\delta T}\right)_P$) is prerequisite to determine most derivatives of thermodynamic properties [43] in HEN model. Peng Robinson Equation of State is selected in this study as it represents the density of the crude oil more accurately in liquid phase due to the consideration of lower critical compressibility factor. Besides, this property package contains enhanced binary interaction parameters for all hydrocarbon-hydrocarbon pairs (a combination of fitted and generated interaction parameters), as well as for most hydrocarbon-nonhydrocarbon pairs. The Peng Robinson Equation of State [43] is given as

$$P = \frac{RT}{V-b} - \frac{a}{V(V+b)+b(V-b)} \tag{3.22}$$

$$a = 0.45723553\frac{R^2 T_c{}^2}{P_c}\left[1+(0.37464+1.54226\omega-0.26992\omega^2)\left[1-\sqrt{T/T_c}\right]\right]$$

$$b = 0.07796074\frac{RT_c}{P_c}$$

where

P = absolute pressure

R = universal gas constant

T = absolute temperature

V = molar volume

T_c = critical temperature

P_c = critical pressure

The compression factor , Z, of Peng Robinson equation is written as

$$f(Z) = Z^3 + (B-1)Z^2 + (A-2B-3B^2)Z + (B^3+B^2-AB) \tag{3.23}$$

$$Z = \frac{PV}{RT}$$

$$\mathrm{A} = \frac{aP}{(RT)^2}$$

$$\mathrm{B} = \frac{bP}{RT}$$

For composition of fluids that have more than one components at specified temperature, the mixture parameters, *a* and *b* are calculated as

$$\mathrm{a} = \sum_{i=1}^{N}\sum_{j=1}^{N} x_i x_j \left(1-k_{ij}\right)\sqrt{a_i a_j} \tag{3.24}$$

$$b = \sum_{i=1}^{N} x_i b_i \tag{3.25}$$

The binary interaction coefficient, k_{ij}, is zero when *i* is equal to *j*. k_{ij} is close to zero for hydrocarbons. Database of k_{ij} are available in the literature [43] The value of mixture parameter *a* is determined based on the known mole fractions of components (mole fraction of component *i*, x_i and mole fraction of component *j*, x_j), parameters *a* for each components (a_i, a_j), and k_{ij} between component *i* and *j*. The value of the mixture parameter *b* is calculated based on the known mole fractions of components, x_i, and parameter *b* for each component, b_i.

The mixture parameters *a* and *b* are used to determine the compressibility factor, *Z*. Newton-Raphson method can be used to solve the Eq. 3.23 numerically to solve for *Z* by considering the largest of the three roots. Consequently, the molar volume of the mixture can be calculated as

$$V = \frac{ZRT}{P} \tag{3.26}$$

With the knowledge of molar volume and compressibility, the three PVT derivatives are determined directly from the equation of state.

These three derivatives must satisfy Cyclical Rule [43] and is written as

$$\left(\frac{\delta P}{\delta V}\right)_T \left(\frac{\delta T}{\delta P}\right)_V \left(\frac{\delta V}{\delta T}\right)_P = -1 \tag{3.27}$$

Therefore, solving of any of the two of the three PVT derivatives, the third derivatives can be determined from Eq. 3.23.

The first derivative is found by direct differentiation of Eq. 3.22.

$$\left(\frac{\delta P}{\delta V}\right)_T = \frac{-RT}{(V-b)^2} + \frac{2a(V+b)}{\left[V(V+b)+b(V-b)\right]^2} \tag{3.28}$$

Similarly, the second derivative can be determined by differentiation of Eq. 3.22. The derivative is given as

$$\left(\frac{\delta T}{\delta P}\right)_V = \frac{R}{V-b} + \frac{a'}{V(V+b)+b(V-b)} \tag{3.29}$$

The third derivative is found implicitly using Eq. 3.21.

$$\left(\frac{\delta V}{\delta T}\right)_P = \frac{R}{P}\left[T\left(\frac{\delta Z}{\delta T}\right)_P + Z\right] \tag{3.30}$$

$$\left(\frac{\delta Z}{\delta T}\right)_P = \frac{\left(\frac{\delta A}{\delta T}\right)_P (B-Z) + \left(\frac{\delta B}{\delta T}\right)_P \left(6BZ + 2Z = 3B^2 - 2B + A - Z^2\right)}{3Z^2\, 2(B-1)Z + (A - 2B - 3B^2)}$$

$$\left(\frac{\delta A}{\delta T}\right)_P = \frac{P}{R^2T^2}\left(a' - \frac{2a}{T}\right)$$

$$\left(\frac{\delta B}{\delta T}\right)_P = \frac{-bP}{RT^2}$$

$$a' = \frac{1}{2}\sum_{i=1}^{N}\sum_{j=1}^{N} w_i w_j (1-k_{ij})\left(\sqrt{\frac{a_j}{a_i}}a_i' + \sqrt{\frac{a_i}{a_i}}a_j'\right)$$

$$a_i' = \frac{-m_i a_i}{\left[1 + m_i(1 - \sqrt{\frac{T}{T_{ci}}}\right]\sqrt{TT_{ci}}}, \quad a_j' = \frac{-m_j a_j}{\left[1 + m_j(1 - \sqrt{\frac{T}{T_{cj}}}\right]\sqrt{TT_{cj}}}$$

Crude oil as produced in the oil field is a complex mixture of hydrocarbons ranging from methane to asphalt, with varying proportions of paraffins, naphthenes, and aromatics [44]. The complete and definitive analysis of a crude oil is called crude assay. A complete crude assay contains a true boiling point (TBP), light-end carbons analysis up to decane, and properties of fractions (naphthas, kerosenes, diesels, heavy diesels, vacuum gas oils, and residues).

The essential physical properties needed in this study are heat capacities, C_p, thermal conductivity, k, specific gravity, SG, and viscosity, μ. For the calculations of physical properties, it is necessary to represent crude oil with discreet components for which thermophysica1 properties may be defined. Since most of the actual components are unknown for crude oil, petroleum pseudo components are developed to represent the unknown components. Pseudo-components represent pre-defined boiling point ranges or cut-point ranges on the true boiling point (TBP) distillation curve for the crude oil that is characterized. Each pseudo-component is corresponding to several unknown actual components. The TBP curve for one of the crude oil used in this study is illustrated as in Figure 3.2.

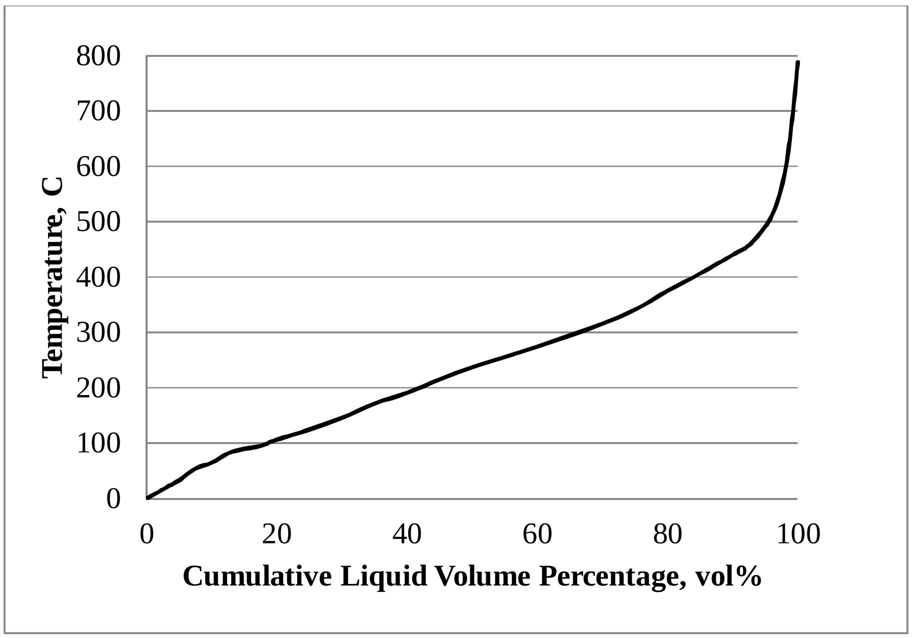

Figure 3.2: TBP curve for one of the crude oil used in this study.

The properties of each pseudo-component include yields as volume percent, molecular weight, gravity, sulfur, viscosity, octane number, diesel index, flash point, fire point, freeze point, smoke point, and pour point. The bulk properties of the pseudo-components are determined based on simple mixing rules [45].

3.2.3.1 Heat Capacity

Liquid heat capacity, C_p, is used to evaluate the thermodynamic properties such as enthalpy. The isobaric heat capacity of crude pseudo-components is determined based on Lee and Kesler approach [45] based on critical region and it is given as

$$C_p = A_1 + A_2 T + A_3 T^2 \;,\; T_r \le 0.85 \tag{3.31}$$

$$A_1 = -1.17126 + (0.023722 + 0.024907SG)K + (1.14982 - 0.046535K)/SG$$

$$A_2 = (10^{-4})(1.0 + 0.82463K)(1.12172 - 0.27634/SG)$$

$$A_3 = (-10^{-8})(1.0 + 0.82463K)(2.9027 - 0.70958/SG)$$

$$K = \frac{T_b^{1/3}}{SG}$$

where

A_1, A_2 , A_3 = characteristic constants of the pseudo-component respectfully

T = operating temperature

T_b = boiling point temperature

K = Watson characterization factor

The Lee and Kesle approach is an industrial benchmark method due to its accuracy in predicting heat capacity of crude containing mainly paraffin and naphthene. It requires only the characteristic constants (A_1, A_2 , A_3) and simple bulk properties such as T_b, SG, and K of the fluid as an input.

3.2.3.2 Density

The SG values at 60°F that are stored in the assay for each pseudo-component are combined based on a volume basis to estimate bulk specific gravity values at 60°F.

$$SG = \frac{\sum_{i=1}^{n} (V_i SG_i)}{\sum V_i} \tag{3.32}$$

Then, API procedure 6A3.6 is used to derive SG at temperature other than 60 °F [45]. The density of liquid petroleum fraction is given as

$$\frac{1}{\rho} = \frac{RT_{pc}}{P_{pc}} Z_{RA}^{\lambda} \tag{3.33}$$

$$\lambda = 1 + (1 - \frac{T}{T_c})^{2/7}$$

where

ρ = density of liquid petroleum fraction

T_{pc} = pseudocritical temperature

P_{pc} = pseudocritical pressure

λ = intermediate variable

Z_{RA} = empirically derived constant

Eq. 3.30 is firstly used to solve for Z_{RA} using the density values at 60°F. The calculated value of Z_{RA} is then used with T reset to the desired temperature, to calculate a value of ρ at T.

3.2.3.3 Thermal Conductivity

The thermal conductivity for liquids is calculated based on API Procedure [1983, 3:12A3.1], with the following curve-fit equation based on mean average boiling point, *MeABP*, in °F [45].

$$k_i = 0.07727 - 4.558x10^{-5} MeABP \tag{3.34}$$

$$MeABP = WABP - 0.795 * (S_{D2887})^{1.83}$$

$$WABP = 0.015(T_{IBP} + T_{FBP}) + 0.051(T_5 + T_{95}) + 0.116(T_{10} + T_{90}) + 0.212(T_{30} + T_{50} + T_{70})$$

where

T_{IBP} = initial boiling point temperature

T_{FBP} = final boiling point temperature

T_i = boiling temperature at i^{th} volume percent of crude assay

3.2.3.4 Viscosity

The viscosity values for hypothetical component, μ_i at different cut points in crude assay are recorded at standard temperatures of 50°C and 100°C [45]. The overall viscosity for mixture of hypothetical components at 50°C and 100°C is determined by combining the viscosity blending number, *VBN*, of pseudo-components as given below.

$$\begin{aligned} VBN_i &= -56 && \text{If } \mu_i \le 0.21 \\ &= 23.1 + 33.47\log_{10}\left[\log_{10}(\mu_i + 0.8)\right] && \text{If } \mu_i > 0.21 \end{aligned} \tag{3.35}$$

The *VBN* for the crude assay (mixture of hypothetical components), VBN^T, is determined as

$$VBN^T = \frac{\sum_{i=1}^{n} V_i^T VBN_i^T}{\sum_{i=1}^{n} V_i^T} \tag{3.36}$$

where

T = temperature of viscosity value either at 50 °C or 100 °C

VBN_i^T = *VBN* of component *i* at temperature *T*

V_i = volume percent of component *i* at temperature *T*

The viscosity of crude assay at either 50 °C or 100 °C, μ_T, is calculated based on equations as given.

$$\begin{aligned} \mu_T &= 0.21 && \text{for } VBN \le -56 \\ &= 10^{10^{\lambda}} - 0.8 && \text{for } VBN > -56 \end{aligned} \tag{3.37}$$

where

$$\lambda = \frac{VBN - 23.1}{33.47}$$

The viscosity values at temperatures other than 50 °C or 100 °C, V_T is determined based on ASTM D341 procedures [45]. The calculations are given as:

$$D = 10^{10^{C}} - 0.7 \tag{3.38}$$

where

$$C = \left(X_1 - \log_{10}(T_1 + 273)\right) + \left(\frac{X_1 - X_2}{\log_{10}(T_1 + 273) - \log_{10}(T_2 + 273)}\right)\log_{10}(T + 273)$$

T_1 = viscosity reference temperature at 50 ^{o}C

T_2 = viscosity reference temperature at 100 ^{o}C

T = temperature at which viscosity is required.

$$\begin{aligned} X_n &= \log_{10}\left\{\log_{10}\left[V_{Tn} + 0.7 + \exp(-1.47 - 1.84V_{T,n} + 0.51V_{Tn}^{\ 2}\right]\right\} && \text{for } 0.11529 < V_{Tn} < 2.0 \\ &= \log_{10}\left\{\log_{10}\left[V_{Tn} + 0.7\right]\right\} && \text{for } V_{TN} \geq 2.0 \end{aligned}$$

where

V_{Tn} = viscosity at temperature T_n with n = 1 or 2

X_n = intermediate values X_n with n = 1 or 2

Based on Eqs. 3.33-3.36, the viscosities of crude oil at 50 ^{o}C and 100 ^{o}C are estimated as given

$$\begin{aligned} V_T &= D - \exp(-0.7487 - 3.295D + 0.6119D^2 - 0.3193D^3 && \text{if } D \leq 2.0 \\ &= D && \text{if } D > 2.0 \end{aligned} \tag{3.39}$$

3.3 Crude Preheat Train Model

The CPT model consists of seven heat exchangers from downstream of desalter till upstream of fired heater. This section of CPT is chosen due to: i) the fouling at upstream of desalter are mainly contributed by unsettled fine particles and minerals which are difficult to estimate using mathematical model, ii) established fouling models [15, 35, 36, 46] are only applicable to chemical reaction fouling, iii) chemical reaction fouling always occurred at elevated temperatures as stated by Wilson et al. [47] and Bott [17]. The mechanical design, physical arrangements, and operational constraints of the heat exchangers are collected from an operating refinery. The schematic diagram of CPT is shown in Fig. 3.3.

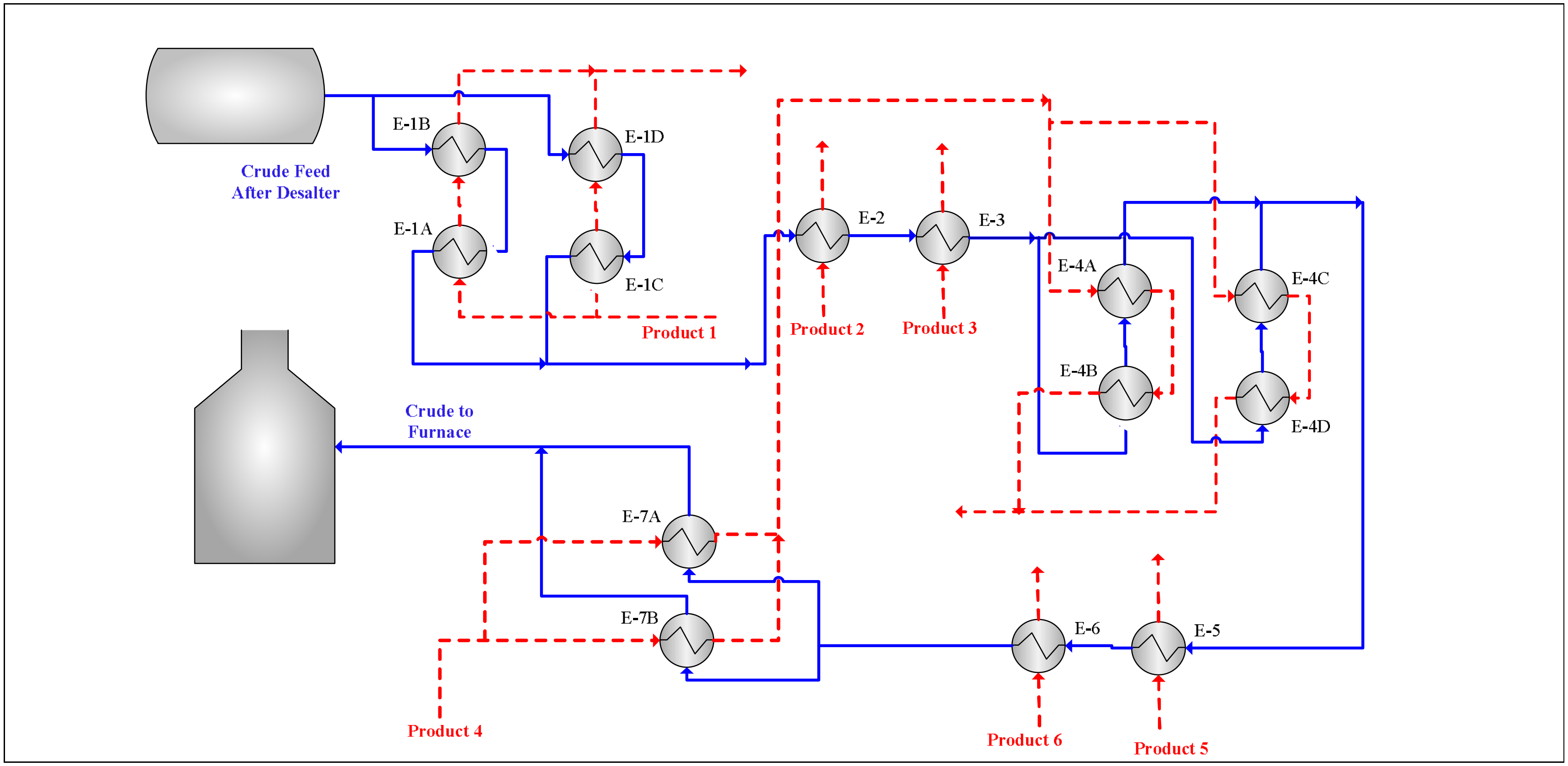

Figure 3.3: Layout of crude preheat train

The crude oil that is free of inorganic salts after desalter is routed through the tube sides of E-1A~D that consists of four shells. The four-shell heat exchanger is formed by two series of heat exchangers, E-1A/B and E-1C/D that are arranged in parallel. After that, the heated crude oil is combined and routed through the tube side of E-2 and subsequently through the tube side of E-3. Both E-2 and E-3 are single-shell heat exchangers. The crude oil is then routed through the shell sides of E-4A~D. E-4A~D has the same shells arrangements as in E-1A~D. The four-shell heat exchanger formed two series of heat exchangers, E-4A/B and E-4C/D that are arranged in parallel. After preheated in E-4A~D, the crude oil is combined and fed to the tube side of E-5 and later through the tube side of E-6. Both E-5 and E-6 are single-shell heat exchangers. Finally, the crude oil is split equally to E-7A and E-7B that are arranged in parallel. The preheated crude oil is then combined and routed through the fired heater to heat to coil outlet temperature (COT) in preparation of separation process in crude distillation column (CDU). The preheating mediums are the hot products and pump-around streams of CDU at different operating conditions. Product 4 has the highest inlet temperature, followed by Product 6, Product 3, Product 2 and Product 1. Product 5 has the lowest inlet temperature. The design specifications of these heat exchangers are summarized in Tables 3.1-3.4.

Table 3.1: Mechanical geometries of heat exchangers

	Heat Transfer Area m^2	Baffle Type	Shell Arrangement Series	Parallel	Tubes per shell	Tube Passes	Velocity, m/s Shell	Tube
1. E-1 A~D	1624	Single Segmental	2	2	91	2	0.50	0.78
2. E-2	238	Single Segmental	1	1	454	1	0.35	2.12
3. E-3	296	Single Segmental	1	1	570	1	0.10	1.70
4. E-4 A~D	544	Single Helix	2	2	99	2	1.83	0.96
5. E-5	495	Single Segmental	1	1	749	1	0.26	1.35
6. E-6	243	Single Segmental	1	1	362	1	0.34	2.89
7. E-7 A/B	295.2	Single Helix	1	2	108	1	1.91	1.20

Table 3.2: Heat Transfer details of heat exchangers

	Heat Duty kW	U W/m^2 ^{o}C	Fluid (Shell)	h_o (Shell) W/m^2 ^{o}C	ΔP (Shell) norm/max kPa	Fluid (Tube)	h_i (Tube) W/m^2 ^{o}C	ΔP (Tube) norm/max kPa
1. E-1 A~D	6985	516	Product 1	1082	29/70	Crude	1034.8	18/50
2. E-2	1629	677	Product 2	714.5	14/70	Crude	2391.8	35/70
3. E-3	1409	175	Product 3	444.1	3/70	Crude	1952.4	24/70
4. E-4 A~D	3280	496	Crude	877.3	60/70	Product 4	944.4	59/70
5. E-5	2854	544	Product 5	500.8	9/70	Crude	1659.4	18/50
6. E-6	6839	699	Product 6	1424.8	14/70	Crude	2868	63/70
7. E-7 A/B	5858	555	Crude	786.9	30/30	Product 4	1334.6	19/70

3.3: Design operations of heat exchangers

	Flow Rates, kg/h		Inlet Temperatures, °C		Outlet Temperatures, °C	
	shell	tube	shell	tube	shell	tube
1. E-1 A~D	377161	420085	159	130	141	147
2. E-2	99246	420085	176	147	157	151
3. E-3	14436	420085	306	151	163	157
4. E-4 A~D	420085	96001	157	256	176	172
5. E-5	121864	420085	265	176	203	195
6. E-6	85828	420085	321	195	242	213
7. E-7 A/B	420085	96001	213	336	232	256

Table 3.4: Fluid physical properties at design operating conditions

	ρ, kg/m^3		Cp, kJ/kgK		μ, cP		k, W/mK	
	shell	tube	shell	tube	shell	tube	shell	tube
1. E-1 A~D	637	716	2.67	2.49	0.202	0.535	0.099	0.093
2. E-2	668	702	2.71	2.57	0.250	0.459	0.090	0.089
3. E-3	678	698	3.24	2.59	0.360	0.443	0.069	0.088
4. E-4 A~D	693	788	2.62	2.76	0.348	1.808	0.087	0.077
5. E-5	640	675	3.05	2.71	0.201	0.348	0.076	0.086
6. E-6	642	656	3.31	2.80	0.227	0.289	0.064	0.089
7. E-7 A/B	637	737	2.89	3.08	0.258	0.377	0.095	0.068

3.3.1 Mathematical Model

The mathematical model of crude preheat train considers serial and parallel arrangements of heat exchangers. The major parameters are fluid temperatures, fluid flow rates and fluid bulk properties (heat capacity (C_p), density (ρ), thermal conductivity (k) and viscosity (μ)). Each of the n heat exchangers is characterized by heat exchange area (A), tube side heat transfer coefficient (h_i), shell side heat transfer coefficient (h_o), and fouling resistance (R_f). These information forms a set of input data to the CPT model.

The heat transfer rate for n^{th} heat exchanger is given by

$$Q_n = U_n A_n F_n . LMTD_n \tag{3.40}$$

Assuming no energy loss, an energy balance on cold and hot streams of n^{th} heat exchanger can be written as

$$Q_c = M_c C_{p,c} (T^c_{n,out} - T^c_{n,in}) \tag{3.41}$$

$$Q_h = M_h C_{p,h} (T^h_{n,in} - T^h_{n,out}) \tag{3.42}$$

$$Q_n = Q_c = Q_h \tag{3.43}$$

Subscripts c and h refer to cold and hot streams, respectively. The overall heat transfer performance, Q of the CPT model is given as

$$Q = \sum_{i=1}^{n} Q_i \tag{3.44}$$

The heat transfer coefficient value from the design data sheet at the design operating conditions is considered to be the base value to determine the actual heat transfer coefficient, h_o. In the calculation, correction factors (J_b, J_l) that are related to mechanical geometries of the heat exchangers are not considered. Correction factor, J_r, is assumed equal to 1 based on the assumption that only turbulent flow exist.

The changes in actual heat transfer coefficients are only due to the variations in the physical properties of the fluids and operating conditions. Therefore the heat transfer coefficients can be calculated as a correction from the design operating conditions and properties. The ratio of actual shell side heat transfer coefficient and shell side design heat transfer coefficients ($h_o/h_{o,d}$) for single segmental baffle is given by :

$$\frac{h_o}{h_{o,d}} = \left(\frac{\rho v}{\rho_d v_d}\right)^{0.65} \left(\frac{\mu}{\mu_d}\right)^{1.1} \left(\frac{C_p}{C_{p,d}}\right)^{2/3} \left(\frac{k}{k_d}\right)^{-2/3} \tag{3.45}$$

Based on Colburn equation, the ratio of actual tube side heat transfer coefficient and design tube side heat transfer coefficient ($h_i/h_{i,d}$) is given by:

$$\frac{h_i}{h_{i,d}} = \left(\frac{\rho v}{\rho_d v_d}\right)^{0.8} \left(\frac{\mu}{\mu_d}\right)^{-0.4} \left(\frac{C_p}{C_{p,d}}\right)^{0.4} \left(\frac{k}{k_d}\right)^{0.6} \tag{3.46}$$

The ratio of actual shell side heat transfer coefficient and shell side design heat transfer coefficient ($h_o/h_{o,d}$) for single helix baffle is given by :

$$\frac{h_o}{h_{o,d}} = \left(\frac{\rho v}{\rho_d v_d}\right)^{0.7} \left(\frac{\mu}{\mu_d}\right)^{-0.4} \left(\frac{C_p}{C_{p,d}}\right)^{1/3} \left(\frac{k}{k_d}\right)^{2/3} \tag{3.47}$$

3.3.2 Fouling Modeling

The fouling resistance, R_f, is given by [42],

$$R_f = \frac{1}{U_t} - \frac{1}{U_{clean}} \tag{3.48}$$

where

U_t = overall heat transfer coefficient at time t

U_{clean} = overall heat transfer coefficient under clean condition

Fouling model as proposed by Panchal et al. [35] is used to determine the fouling rates of heat exchangers and is given by

$$\frac{dR_f}{dt} = \alpha \operatorname{Re}^{-\beta} \operatorname{Pr}^{-0.33} \exp\left(\frac{-E}{RT_f}\right) - \gamma \tau_W \tag{3.49}$$

where

$$\tau_w = \frac{1}{2} f \rho v^2$$

$$f = 0.0035 + \frac{0.264}{\operatorname{Re}^{0.42}} \qquad \text{for turbulent flow}$$

$$T_f = T_b + 0.55(T_w + T_b)$$

α = fouling deposition constant

γ = fouling removal constant

E = activation energy of crude

T_b = bulk temperature of heat exchanger

T_w = wall temperature of heat exchanger

T_f = film temperature of heat exchanger

τ_w = shear stress

β = constant

A value of 0.88 for β is recommended [15] to determine an effective film thickness for the thermal-boundary layer. The equation is then expressed as

$$\frac{dR_f}{dt} = \alpha \operatorname{Re}^{-0.88} \operatorname{Pr}^{-0.33} \exp\left(\frac{-E}{RT_f}\right) - \gamma \left(\frac{0.0791 \rho v^2}{2 \operatorname{Re}^{1/4}}\right) \tag{3.50}$$

Integrating the Eq. 3.48, the cumulative R_f for n^{th} heat exchanger is given by

$$R_{f_{i+1}} = R_{f_i} + \int_{t_i}^{t_{i+1}} \alpha \operatorname{Re}^{-0.88} \operatorname{Pr}^{-0.33} \exp\left(\frac{-E}{RT_f}\right) - \gamma \left(\frac{0.03955 \rho v^2}{\operatorname{Re}^{1/4}}\right) dt \tag{3.51}$$

In this study, fouling model parameters are extracted from historical data based on the assumptions that: i) fouling mechanisms in the selected case study involved chemical reaction fouling, and ii) the fouling happens only at tube side of heat exchangers in CPT.

R_f for heat exchangers in CPT are calculated from the historical data of an operating refinery. Then, the fouling rate, $\frac{dR_f}{dt}$, is determined as the slope of the graph, R_f versus time for the selected time frames. The assumptions for selection of time frames are: (i) same crude oil blend is processed throughout the selected time frame, (ii) drop in R_f profile due to mechanical cleaning or hot melting of heat exchangers will not be included, and (iii) outliers of operating parameters will not be included. τ_w for each heat exchanger is calculated based on average values of calculated *Re, Pr* and T_f during the selected duration. $\frac{dR_f}{dt}$ versus τ_w is then plotted for each heat exchanger to determine γ that is equivalent to the slope of the graph. Then the α and *E* are solved simultaneously using Microsoft Excel for minimum root mean square error, *RMSE,* by comparing the differences of calculated $\frac{dR_f}{dt}$ versus actual $\frac{dR_f}{dt}$ derived from actual operating data.

For E-1A~D, E-3 and E-5 that have approximately the same shear stress, the activation energy, *E*, of the crude oil is computed as average value of the results of E-1A~D, E-3 and E-5.

For E-2 and E-6 that have different τ_w, a different approach is used to determine α and γ. The average *E* that is estimated from simultaneous convergence of E-1A~D, E-3 and E-5 is assumed valid for E-2 and E-6 as fouling is assumed to be occurred at crude oil side. Then, α and γ are solved to determine the minimum *RMSE,* by comparing the differences of calculated $\frac{dR_f}{dt}$ versus actual $\frac{dR_f}{dt}$ that derived from actual operating data.

For E-4A~D and E-7A/B that have different heat transfer medium (Product 4) through the tube side, the fouling model parameters, α, γ and *E*, are solved simultaneously using Microsoft Excel for minimum root mean square error, *RMSE,* by comparing the differences of calculated $\frac{dR_f}{dt}$ versus actual $\frac{dR_f}{dt}$ derived from actual operating data.

3.4 Summary

This chapter presented the development of CPT simulation model in detail. The governing equations, estimation of overall heat transfer coefficients and established first principle thermodynamic correlations for physical properties used in the mathematical model were discussed. The mechanical arrangements, geometries and design specifications of the CPT model were presented. Lastly, the fouling model and the approaches to estimate fouling model parameters were also presented.

CHAPTER 4

SIMULATION STUDIES

4.1 Introduction

A good understanding of effects of fouling on individual heat exchangers in a heat exchanger network (HEN) is important to prioritize the cleaning process in minimizing the operating cost. The analysis can be performed by studying the heat transfer interactions between the heat exchangers in terms of physical arrangements, positions, thermal designs and fouling.

In this section, a simulation study was performed using a mathematical model of crude preheat train (CPT) to analyze the effects of fouling on each heat exchanger and overall CPT. Semi-empirical fouling model is used to estimate the fouling behavior, with the model parameters extracted from historical data collected from a refinery that processes sweet crude oil. The details and results of simulation study are discussed in this section.

4.2 Simulation of the CPT Under Fouled Conditions

Simulation of CPT under fouled conditions comprises three essential elements, which are; i) input data (Temperature, pressure and mass flow) to the CPT model, ii) fouling parameters (deposition constant, α, and removal constant, γ) for each heat exchanger, and iii) mathematical model of CPT under Petro-SIM™ environment. Details of input data and fouling parameters are discussed in this section. The details of CPT mathematical model are discussed in Section 3.3.

4.2.1 Input Data

The input data to the CPT model was extracted from the historical operating data collected from a refinery which processes sweet crude oils. In this simulation analysis, the operating data for the process inlet conditions to the simulation model are summarized in Table 4.1. These input data are kept constant throughout the simulation period. Product streams 1-6 are mainly the products from crude distillation column (CDU) that were routed through different heat exchangers in the CPT to preheat crude A. The pressure drop for all the heat exchangers in the CPT are assumed at normal operating conditions, as indicated in Table 4.1 and it is assumed that the increase in pressure drops in each heat exchanger due to fouling does not affect the fluid flow hydraulics. The heat transfer details of the heat exchangers are listed in Table 4.2. Table 4.3 summarizes the physical properties of fluids at clean operating conditions.

Table 4.1: The input operating data for the simulation model

	Inlet Stream	Inlet Temperature ^{o}C	Inlet Pressure kPa	Flow Rate kg/h
1	Product 1	174.74	885.9	257873
2	Product 2	208.74	1425	38856
3	Product 3	265.94	268	15051
4	Product 4	287.39	1259	81546
5	Product 5	176.27	748.6	106032
6	Product 6	324.11	513.2	108440
7	Crude A	118.54	2359	418233

Table 4.2: Heat transfer details of heat exchangers

	Heat Duty kW	U $W/m^2\ ^oC$	Fluid (Shell)	h_o (Shell) $W/m^2\ ^oC$	ΔP (Shell) norm kPa	Fluid (Tube)	h_i (Tube) $W/m^2\ ^oC$	ΔP (Tube) norm kPa
1. E-1 A~D	6985	516	Product 1	1082	29	Crude	1034.8	18
2. E-2	1629	677	Product 2	714.5	14	Crude	2391.8	35
3. E-3	1409	175	Product 3	444.1	3	Crude	1952.4	24
4. E-4 A~D	3280	496	Crude	877.3	60	Product 4	944.4	59
5. E-5	2854	544	Product 5	500.8	9	Crude	1659.4	18
6. E-6	6839	699	Product 6	1424.8	14	Crude	2868	63
7. E-7 A/B	5858	555	Crude	786.9	30	Product 4	1334.6	19

Table 4.3: Fluid physical properties at clean operating conditions

	ρ kg/m3		C_p kJ/kgK		μ cP		K W/mK	
	shell	tube	shell	tube	shell	tube	shell	tube
1. E-1 A~D	637	716	2.67	2.49	0.202	0.535	0.099	0.093
2. E-2	668	702	2.71	2.57	0.250	0.459	0.090	0.089
3. E-3	678	698	3.24	2.59	0.360	0.443	0.069	0.088
4. E-4 A~D	693	788	2.62	2.76	0.348	1.808	0.087	0.077
5. E-5	640	675	3.05	2.71	0.201	0.348	0.076	0.086
6. E-6	642	656	3.31	2.80	0.227	0.289	0.064	0.089
7. E-7 A/B	637	737	2.89	3.08	0.258	0.377	0.095	0.068

4.2.2 Fouling Parameters

The fouling model proposed by Panchal et al. [35] is chosen to describe the fouling rates of heat exchangers. The fouling rates depend on the operating conditions across the CPT. The fouling model is given by

$$\frac{dR_f}{dt} = \alpha \, \mathrm{Re}^{-\beta} \, \mathrm{Pr}^{-0.33} \exp\left(\frac{-E}{RT_f}\right) - \gamma \tau_W \tag{3.24}$$

The fouling model parameters (deposition constant, α, and removal constant, γ) are estimated from historical data as discussed in Section 3.3.2. These fouling parameters for all the heat exchangers are summarized in Table 4.4.

Table 4.4: Fouling model parameters for heat exchangers in the CPT based on actual operating data.

	Heat Exchangers	Deposition constant, α	Removal constant, γ	Activation Energy, E, kJ/mol
1	E-1A~D	1.031E+05	1.227E-05	39.783
2	E-2	1.200E+06	5.630E-05	39.783
3	E-3	1.031E+05	1.227E-05	39.783
4	E-4A~D	1.000E+07	1.009E-04	52.515
5	E-5	1.031E+05	1.227E-05	39.783
6	E-6	8.860E+05	7.520E-05	39.783
7	E-7A/B	1.000E+07	1.009E-04	52.515

The fouling model parameters were extracted based on assumptions that fouling of heat exchangers in CPT is only occurring at tube side. As indicated in Table 4.4, all the heat exchangers in the CPT have unique combinations of α and γ. E-1A~D, E-3, and E-5 have the same fouling model parameters as these heat exchangers process the same heat transfer fluid at tube side. E-2 and E-6 have different fouling parameters due to different shear stress although its tube side heat transfer fluid is crude oil. E-4A~D and E-7A/B have the same fouling model parameters as the same heat transfer fluid flows through the tube side. As discussed by Panchal et al. [48], fouling model parameters should be consistent across the HEN that shared the same heat transfer medium at tube side. In other words, only chemical reaction fouling is assumed in

Panchal's case study. Similar approach is used in this study as all the fouling model parameters were extracted from historical data collected from refinery, in which the fouling is assumed solely contributed by chemical reaction fouling.

Activation energy, E, for the crude oil that flows through the tube side of E-1A~D, E-2, E-3, E-5 and E-6 have been estimated to be 39.783 kJ/mol, while the E for Product 4 that flows through the tube side of E-4A~D and E-7A/B was 52.515 kJ/mol. All these ranges of E are within reported ranges of activation energy. Crittenden et al. [49] reported the E for crude oils in the range of 20 and 55 kJ/mol.

4.3 Results and Discussion

The CPT model was simulated for a period of 300 days. Throughout the simulation period, the changes in physical properties such as density, heat capacity, viscosity and thermal conductivity of crude oil and preheating mediums for intermediate streams of each heat exchanger were determined by Petro-SIM™ software. Then, the overall heat transfer coefficient is determined by

$$\frac{1}{U} = \frac{d_o}{d_i h_i} + \frac{d_o R_i}{d_i} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2k_w} + R_o + \frac{1}{h_o} \tag{3.7}$$

The shell side heat transfer coefficient, h_o, was calculated using Bell-Delaware method for vertical segmental baffles and Peng's method for continuous helical baffles while the tube side heat transfer coefficient, h_i, were calculated using Colburn method, as explained in Chapter 3, Section 3.2.2.1. Fouling model for each heat exchanger in the CPT was then used to compute the fouling resistance for each day of the simulation period. The intermediate operating temperatures of heat exchanger vary due to the incremental fouling resistance in each heat exchanger estimated by the fouling model. The fouling resistances for all the heat exchangers are shown in Fig. 4.1.

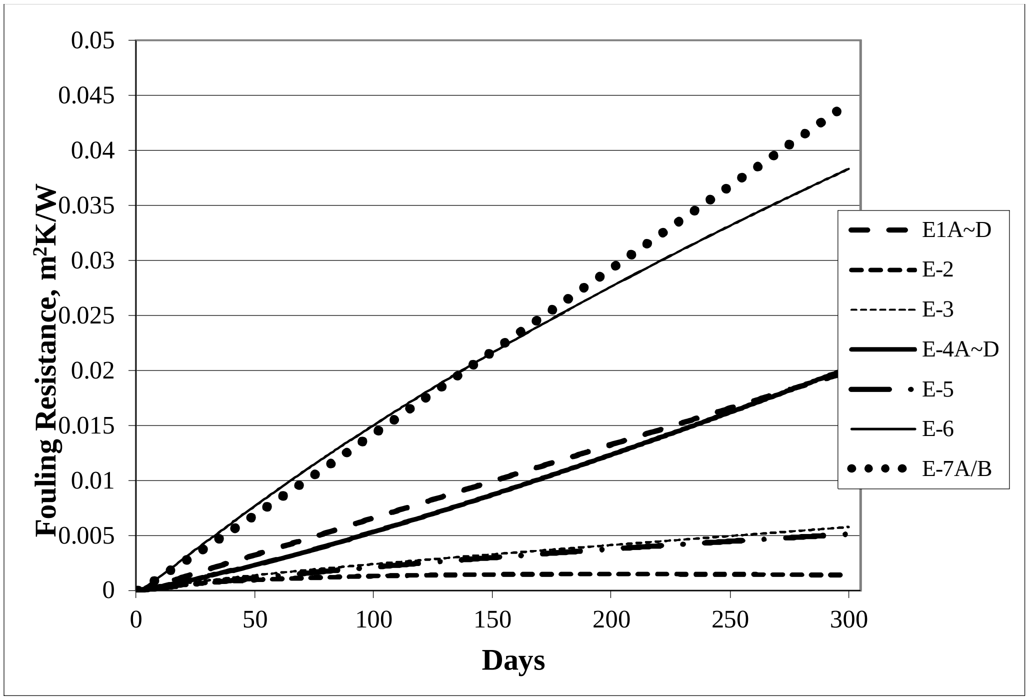

Figure 4.1: Fouling resistances determined for all the heat exchangers.

It is observed that the fouling rates are high for some heat exchangers while it is low for other heat exchangers. The two variables that are responsible for different fouling rates in different heat exchangers are the fluid velocity and the wall/film temperature. These two operating parameters vary across the heat exchangers in the CPT depending on the mechanical geometries and operating conditions of the heat exchangers. The velocity and film temperature data for all the heat exchangers in the CPT are summarized in Table 4.5.These two operating parameters vary across the heat exchangers in

Table 4.5: Velocity and wall/film temperatures for heat exchangers in the CPT based on design operating data

	Heat Exchangers	velocity, v	Wall/film temperature, K
1	E-1A~D	0.708	406.5
2	E-2	2.145	417.7
3	E-3	1.521	423.6
4	E-4A~D	0.467	492.8
5	E-5	1.368	462.6
6	E-6	2.953	490.0
7	E-7A/B	1.290	554.8

Refering to Eq. 3.5, the deposition term is inversely proportional to $v^{0.88}$ and directly proportional to the Arrhenius term, $\exp\left[-\frac{E}{RT_f}\right]$, while the removal term is directly proportional to $v^{1.75}$.

Referring to Table 4.5, the effect of T_f is more dominant as compared to effect of v. Heat exchangers E6 and E-7A/B that operates at the highest T_f have the highest fouling rates. E-1A~D and E-4A~D have relatively lower fouling rates as compared to E6 and E-7A/B. E-3 and E-5 have approximately the same fouling rates, which is lower than E-1A~D and E-4A~D. Heat exchanger E-2 has the lowest fouling rate throughout the simulation period as its high fluid velocity slow down the fouling process. It is observed that the fouling rate in E-2 is asymptotic. Starting from 60th to 110th day, lower fouling rates have been predicted for E-2.

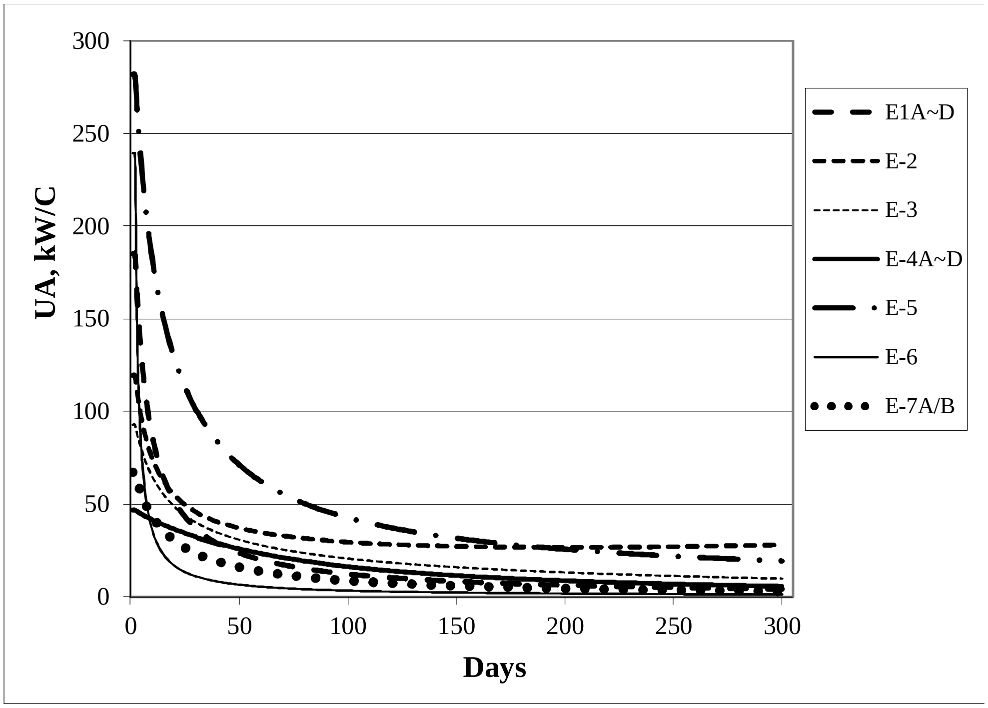

Figure 4.2: *UA* values determined for all the heat exchangers.

The overall heat transfer coefficients for all the heat exchangers were calculated for every day based on the corresponding fouling resistances. The *UA* values for all the heat exchangers are shown in Fig. 4.2. It is observed from Fig. 4.2 that the decrease in *UA* values correspond to the fouling rates in the respective heat exchanger. It is observed from Fig. 4.2 that the *UA* values drop significantly in the first few days of simulation period and reach almost constant values on the longer run. It may be noted that although the R_f values were increasing linearly, the *UA* values asymptotically reached a constant value.

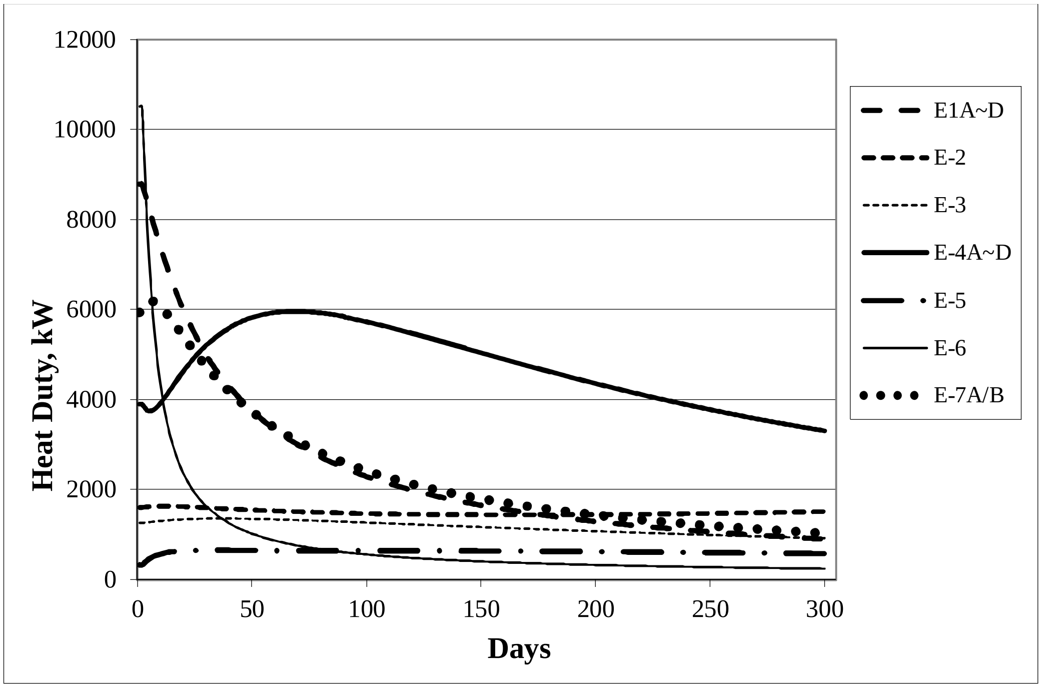

Figure 4.3: Heat duty values determined for all the heat exchangers.

Fig. 4.3 shows the variations in the heat duty in individual heat exchangers due to fouling during the simulation period. It is observed that heat duties are decreasing for most of the heat exchangers in the CPT except E-7A/B, E-3, E-4A~D and E-5. The decrease of heat duties for most of the heat exchangers in the CPT are due to fouling. Higher heat transfer loss is indicated on E-6 (10,003kW) and followed by E-1 (6,666kW). A slight decrease of heat duty is observed in E-2 (139kW). It is observed that in some heat exchangers, e.g. E4-A~D, the heat duty initially increase despite of the fouling and decreased *UA* values. This is mainly due to the increase in the approach temperatures that resulted in higher *LMTD* values.

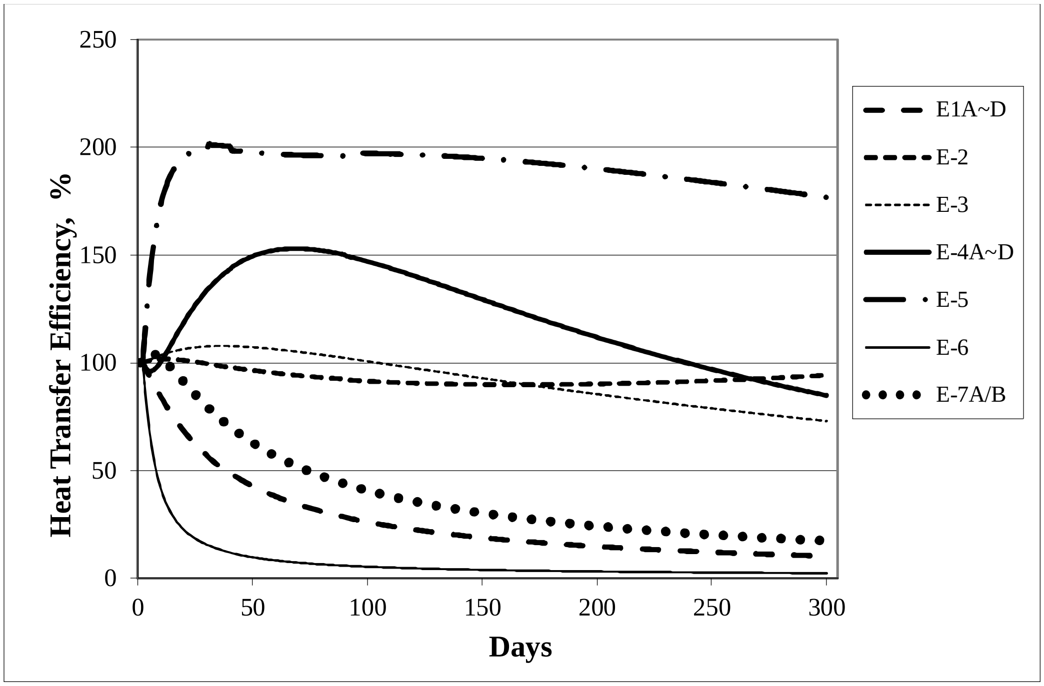

Figure 4.4: Heat transfer efficiency determined for all the heat exchangers.

Fig. 4.4 shows the heat transfer efficiency in individual heat exchangers undergoing fouling during the simulation period. As observed in Fig. 4.4, heat transfer efficiency of most of the heat exchangers dropped due to fouling in asymptotic manner except E-4A~D, E-5 and E-3. Heat exchanger E-6 has the fastest decreasing rate of heat transfer efficiency as it has one of the highest fouling rates as indicated in Fig. 4.1. The high fouling rate caused its heat transfer efficiency to drop to approximately 20% in the first 20 days. The heat transfer efficiency of E-2 decrease for the first 120 and increased slightly till the end of the simulation period. During the first 120 days, the fouling in E-1 caused a drop in the heat transfer efficiency of E-1 and E-2. This observation is aligned with the fouling characteristics of E-1 and E-2. At the end of 120 days, heat transfer efficiency of E-1 dropped to approximately 25%, and this caused an increase in the inlet temperature of the process fluid to E-2. With constant inlet temperature of the heating medium to E-2, the approach temperature increased leading to higher LMTD values. Small variation in the heat transfer efficiency was observed in E-2, whereas its heat transfer efficiency remained around 90-110 % despite the increase in the fouling resistance throughout the simulation period.

Heat transfer efficiency of E-4A~D increased slightly during the first 50 days and dropped after that. The heat transfer performance of E-5 improved drastically up to 200% within the first 40 simulation days. As compared to E-5, the heat transfer performance of E-4A~D improved only up to 150% within the first 75 days.

As observed in Fig. 4.2, fouling of CPT have the least impact to the UA values of E-7A/B and E-4A~D. The heat transfer performance of E-4A~D is highly dependent on E-7A/B as preheating medium of E-4A~D is routed from the outlets of E-7A/B. As indicated in Fig. 4.4, an increase in heat transfer efficiency is observed in E-7A/B during the initial 10 days due to the combined effect of low fouling and high *LMTD* value. Low fouling will contribute to higher *U* while high *LMTD* will results a higher heat transfer across the heat exchanger. In the same duration, the heat transfer performance of E-4A~D was observed to decrease as the inlet temperature of preheating medium from the outlets of E-7A/B are hotter, which reduces the *LMTD* for E-4A~D. Hence, less duty transfer is observed for E-4A~D. Colder crude oil at outlets of E-4A~D increases the *LMTD* of E-5 and hence improved the heat transfer by 37%. This is supported by Fig. 4.4 that indicated the heat transfer efficiency of E-5 increased to 200% during this period.

After 10 days, higher fouling in E-7A/B has resulted in lower LMTD values, and hence reduced the heat transfer performance of E-7A/B. This can be observed from Fig. 4.3, which the heat transfer for E-7A/B is increasing at initial stage of the simulation time, and decrease after few days. Due to less heat exchange in E-7A/B while the same operating temperature and flow of preheating medium are maintained, hotter preheating medium is routed to E-4A~D. The hotter preheating medium increased the *LMTD* and compensated the decrease of *U* of E-4A~D. This LMTD effect increased the heat transfer duty of E-4A~D by 1.71 MW, which is approximately equivalent to a performance improvement of 44%.

Comparison of reduction in heat duties in the individual heat exchangers with the total heat duty loss due to fouling shows that E-6 and E-1A~D results in the highest fouling impact to the total duty loss in the CPT. As observed from Fig. 4.3, E-6 and E-1A~D contributed to 30.9% and 20.6% of the total heat loss of the CPT respectively due to effects of fouling and LMTD.

High fouling of E-7A/B has increased the crude oil outlet temperature of E-4A~D. This hence decreases the heat transfer driving force for the heat exchangers at the downstream of CPT. E-6 has the most drastic decrease in heat transfer as it has the highest fouling tendency among all the heat exchangers in the CPT as observed in Fig. 4.2. Lowest *LMTD* effect is observed in E-1A~D. All the operating temperatures and flows to this heat exchanger is fixed as this is the first heat exchanger of the CPT. The changes of *LMTD* values are merely due to changes of intermediate outlet temperatures of E-1A to E-1B that are arranged in series. Similar effect is observed on E-1C and E-1D.

The total heat duty for the CPT for the simulation period is shown in Figure 4.5. The total heat recovery of the CPT had dropped by 23.99 MW from 32.33 MW to 8.33 MW at the end of 300 simulation days due to fouling.

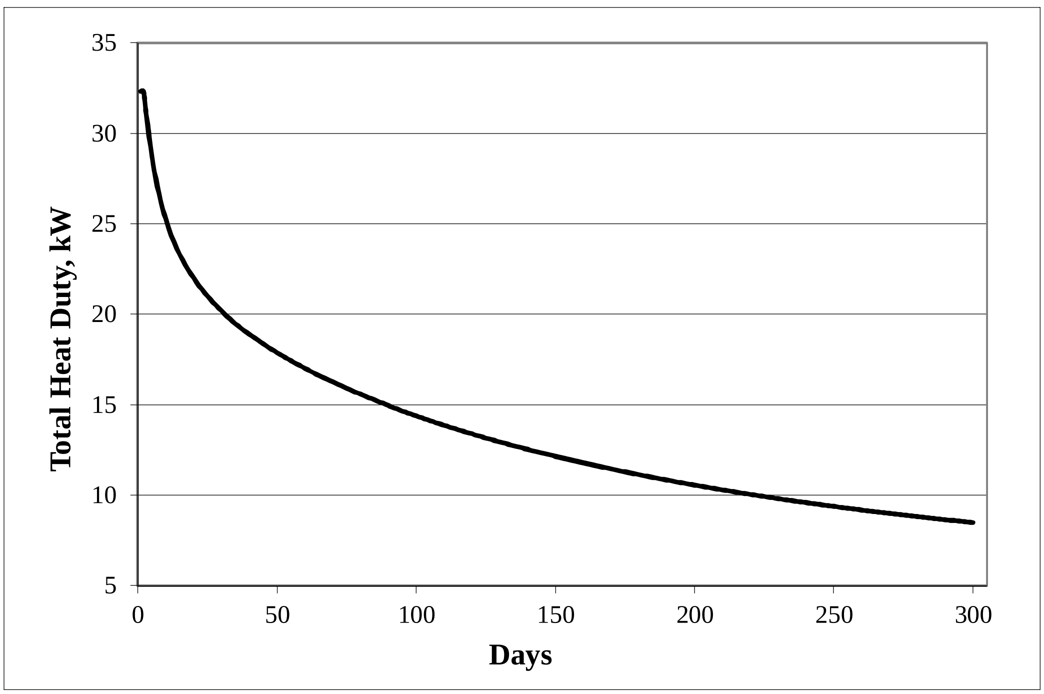

Figure 4.5: Total heat transfer value determined for all the heat exchangers.

The crude oil from the last heat exchanger in the CPT is pumped into the fired heater to further heat the crude oil to the operating feed temperature of CDU. The inlet temperature to the fired heater is called the coil inlet temperature (CIT).

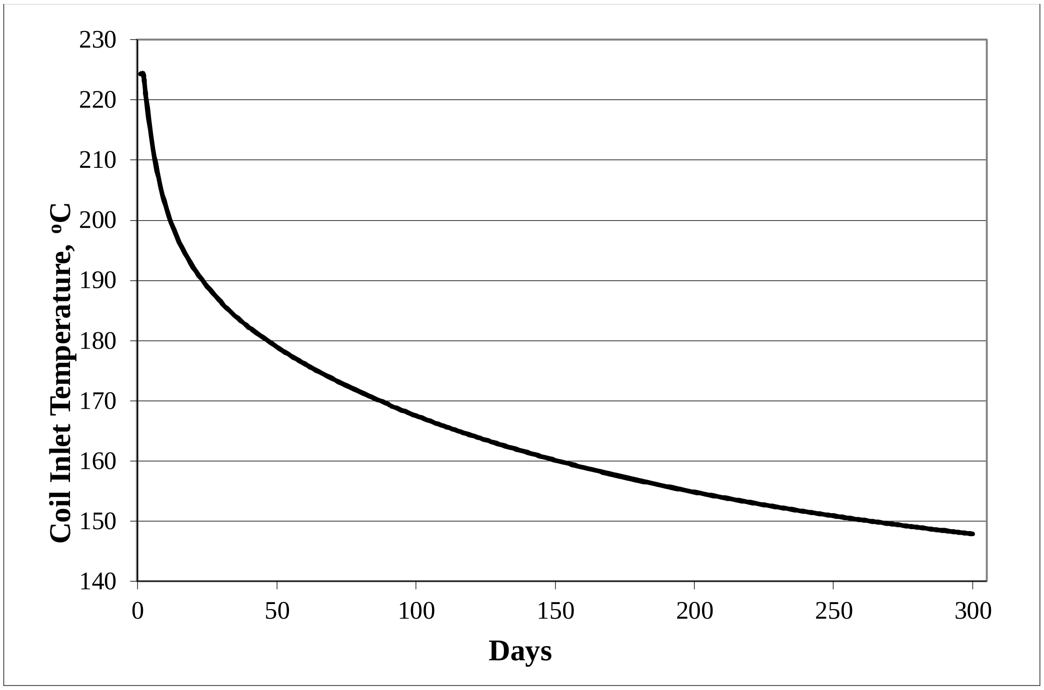

Figure 4.6: CIT values determined for all the heat exchangers.

Fig. 4.6 shows that coil inlet temperature (CIT) dropped by 76.93 °C from 224.3 to 147.39 °C at the end of 300 days. Based on the calculation for a preheat train processing similar throughput as provided by refinery, the cost of fuel gas is approximately RM11.30 to generate one Giga Joule of heating energy in fired heater as reported from an operating refinery. Thus, the reduction in CIT results in additional fuel cost equivalent to RM 8.72 mil in one year.

4.4 Summary

This chapter presented a heat transfer analysis of the seven units of heat exchangers after desalter in the CPT for a crude distillation unit in 300 days. Same operating conditions were input to inlet process boundaries of the CPT model under Petro-SIM™ environment throughout the simulation duration. Fouling model as proposed by Panchal et. al. [15] was used to estimate the fouling rates of heat exchanger, with its fouling parameters were estimated from historical operating data from an operating refinery.

Different heat transfer performances of heat exchangers are observed during the simulation period due to fouling. Heat transfer performance dropped for some of the heat exchangers and increased for other heat exchangers. Drop in heat transfer performances are observed on E-6 and followed by E-1 and E-2 due to fouling. Improvement on heat transfer performances are observed on E-7A/B, E-3, E-4A~D and E-5. These improvements are mainly contributed by *LMTD* effects in CPT.

The analysis indicated that position of heat exchangers has essential roles in heat transfer performance in the CPT during fouling condition. As indicated in this simulation study, fouling of first heat exchanger of the CPT will have higher impact to overall heat transfer performance of the CPT. In this study, it is observed that even at consistent increase of fouling rates, the loss of heat performance due to fouling can be compensated by *LMTD* effect. To certain extent, *LMTD* effect will slow down fouling effect or even increase the heat transfer performance in some of the heat exchangers in the HEN.

CHAPTER 5

OPTIMIZATION OF CLEANING SCHEDULE

5.1 Introduction

Cleaning of heat exchangers to remove deposits due to fouling in crude preheat train (CPT) is a normal ad-hoc operation in oil refineries. The performance of CPT is evaluated by the plant operators solely based on operating experience. In general, plant operators have the views that the heat exchangers should be cleaned when the CPT exceeds the operational limits. These operational limits are abnormally high pressure drops across the heat exchangers, insufficient firing in the fired furnace to achieve targeted coil outlet temperature (COT) at full plant throughput, and production schedule from planning department. The ad-hoc and improper planning of cleaning schedule of CPT will affect the productions of the refinery and hence result in economic losses. This chapter discusses the formulation and solution of an optimization problem for the cleaning schedule of CPT.

A base case for the cleaning of heat exchangers has been developed based on the maximum allowable fouling resistances in the heat exchangers that correspond to maximum allowable pressure drops across the heat exchangers. Operations beyond the maximum allowable pressure drops lead to reduction in the plant throughput to overcome the high pressure drops by the feeding pumps.

This ad-hoc cleaning generally results in: i) higher economic loss due to unrecovered energy, ii) unplanned reduced throughputs with high penalties, and iii) operating difficulties. An optimal cleaning schedule taking into account all the factors will result in a minimal operating cost with less disruption in the plant operations.

5.2 Formulation of Optimization Problem

The model developed to describe the CPT, as explained in Section 3.3, is used in the study of optimization of cleaning schedule of the CPT.

5.2.1 Objective Function

An optimal cleaning schedule of the CPT depends on the following costs:

i. Unrecovered energy cost: The unrecovered energy is the difference between the heat transfer duties at fouled conditions and the corresponding clean conditions in the heat exchangers in the CPT.

ii. Cleaning costs: The cleaning costs of heat exchangers involve the costs associated with the initial preparations for heat exchanger isolation, mechanical/chemical cleaning and reinstallation of the heat exchangers. The cleaning costs generally depend on the size of the heat exchangers.

Based on the above costs, an objective function is formulated as shown in Eq. 5.1.

$$\min_{y_{n,t}} J = \sum_{t=1}^{t_p} \sum_{n=1}^{N_E} \left[C_E \, y_{n,t} (Q_{n,clean} - Q_{n,t}) + C_{cl} (1 - y_{n,t}) \right] \tag{5.1}$$

where

t_p = total period of operation

t = time, day

n = n^{th} heat exchanger

C_E = cost of energy per day

$Q_{n,clean}$ = heat transfer duty under clean conditions in the n^{th} heat exchanger

$Q_{n,t}$ = heat transfer duty under fouled (actual) conditions in the n^{th} heat exchanger

C_{cl} = cost of heat exchanger cleaning per day

N_E = the total number of heat exchangers in the CPT

$y_{n,t}$ = the cleaning status of n^{th} heat exchanger unit at time t

The cleaning status of the heat exchangers, $y_{n,t}$, is a binary variable, where it is equal to 1 if n^{th} heat exchanger is in service, and it is equal to 0 if it is under cleaning.

In the optimization problem of Eq. 5.1, the optimization variables are the cleaning status of the heat exchangers. The number of optimization variables is equal to the total optimization period in days for each heat exchanger, which is extremely large in number. In practice, the cleaning status of any heat exchanger is equal to 1 for the whole duration of the heat exchanger between successive cleanings. Therefore, the cleaning status of heat exchangers can be replaced with the duration between successive cleaning which results in a fewer number of optimization variables. In this study, the duration between cleanings, $t_{s,n}$, in days will be the optimization variables. The optimization problem needs to be solved using mixed integer programming approach to determine the optimal duration between cleanings of heat exchangers in the CPT.

The starting day of the j^{th} cleaning for the n^{th} heat exchanger in the optimization cases, is determined by

$$t_{n,j} = t_{s,n} j_n + t_{c,n}(j_n - 1) + 1 \tag{5.2}$$

where

$t_{c,n}$ = cleaning duration for n^{th} heat exchanger

j_n = j^{th} cleaning cycle of n^{th} heat exchanger

All the heat exchangers are assumed that they can be bypassed either independently or sharing the same bypass facility with another heat exchanger. Heat exchangers E-1A~D, E-4A~D and E-7A/B can be bypassed independently. E-2 and E-3 share the same bypass facility. Same bypass facility is also equipped for E-5 and E-6. The heat exchangers that share the same bypass facility must be cleaned simultaneously during the cleaning period regardless of the fouling condition of other heat exchanger that shares the bypass facility.

The cleaning of the seven heat exchangers in the CPT model depends on the arrangement of shells, i.e., either in series or in parallel. For E-1A~D and E-4A~D that have 2 trains of heat exchangers in parallel, cleaning of one heat exchanger train will not affect the operation of the other heat exchanger train. Heat exchanger E-7A/B consists of 2 heat exchangers in parallel. Similar to cleaning cycles of E-1A~D and E-4A~D, cleaning of one heat exchanger will not affect the operation of the other heat exchanger. The cleaning of the second heat exchanger is performed after cleaning the first heat exchanger.

The cleaning duration for one heat exchanger is assumed to be 5 days, based on the industrial experience. Same cleaning duration is applied for one heat exchanger train that consists of two heat exchangers, assuming both the heat exchangers are cleaned simultaneously. The cleaning of second heat exchanger train is performed after the cleaning cycle of first train. With that, the cleaning durations for E-1A~D, E-4A~D and E-7A/B are 10 days for each cleaning. The cleaning durations for E-2, E-3, E-5 and E-6 are 5 days for each cleaning.

The amount of supplement heating energy required to maintain coil inlet temperature (CIT) at 225 °C was measured at the outlet of heat exchanger E-7 A/B. The supplemented heating energy is then converted to monetary basis based on the value of energy cost, C_E , which is equivalent to RM11.30/GJ.

The cleaning costs of heat exchangers differ based on the heat transfer area, mechanical arrangements and availability of bypass facilities. The nominal cleaning costs for the heat exchangers in the CPT are obtained from the plant as shown in Table 5.1.

Table 5.1: Cleaning durations and mechanical cleaning costs of heat exchangers in CPT.

	Heat Exchangers	$t_{c,n}$ (days)	$C_{cl,ni}$ (RM/day)
1.	E-1A~D	10	30, 000
2.	E-2	5	10, 000
3.	E-3	5	10, 000
4.	E-4A~D	10	20, 800
5.	E-5	5	11, 800
6.	E-6	5	9, 600
7.	E-7A/B	10	18, 300

Fouling results in the formation of roughened deposits adhering to the surface of the heat exchanger tubes. The crude oil is often passed through the tubes and the flow area reduces as these deposits grow, increasing the flow resistance encountered within the exchanger and ultimately the network. There are occasions when the flow resistance becomes so high that the operators have to reduce throughput. These flow resistances can be measured based on pressure drops of the heat exchangers and these parameters are crucial to be monitored in this optimization study. In general, the pressure drops in the heat exchangers are directly proportional to the corresponding fouling resistances [22]. An increase in the fouling deposition in the heat exchangers increases the pressure drop, which reduces the throughput across the heat exchangers.

Nevertheless, changes in pressure drop for each heat exchanger due to the variations in throughput are not simulated in this optimization study. Alternatively, the highest allowable fouling resistances for heat exchangers extracted from historical data are used as one of the constraints. The fouling resistance constraints for all the heat exchanger are indicated in Table 5.2.

Table 5.2: Maximum allowable fouling resistances for the heat exchangers in CPT

	Heat Exchanger	Fouling Resistance W/m^2K
1.	E-1A~D	2.30E-02
2.	E-2	1.06E-02
3.	E-3	1.30E-02
4.	E-4A~D	1.44E-02
5.	E-5	1.49E-02
6.	E-6	2.30E-02
7.	E-7A/B	1.80E-02

5.2.2 Constant Fouling Rates

Different types of crude oils were blended and preheated in the CPT during the period for which the historical operating data was collected from the plant. Mostly, a crude oil blend is only processed for less than two weeks and none of the crude oil blends were repeatedly processed for more than three times in the two years duration.

The fouling rates of each heat exchanger in the CPT determined from the historical data were plotted in Figs. 5.1-5.7. Equal fouling rate is assumed in all the shells of a multi-shell heat exchanger. It is observed from Figs. 5.1-5.7 that the fouling resistances increased nearly linearly in all the heat exchangers. The fouling rates at different points in time in any heat exchanger are almost the same even though various types of crude oil blends were processed at different times. Thus, it is assumed in this optimization study that all the heat exchangers have constant linear fouling rates. The observed fluctuations in the values of fouling rates were mainly due to variations in the crude oil blends and also the velocity due to the variations in the throughput. The constant fouling rates for heat exchangers in the CPT are summarized in Table 5.3.

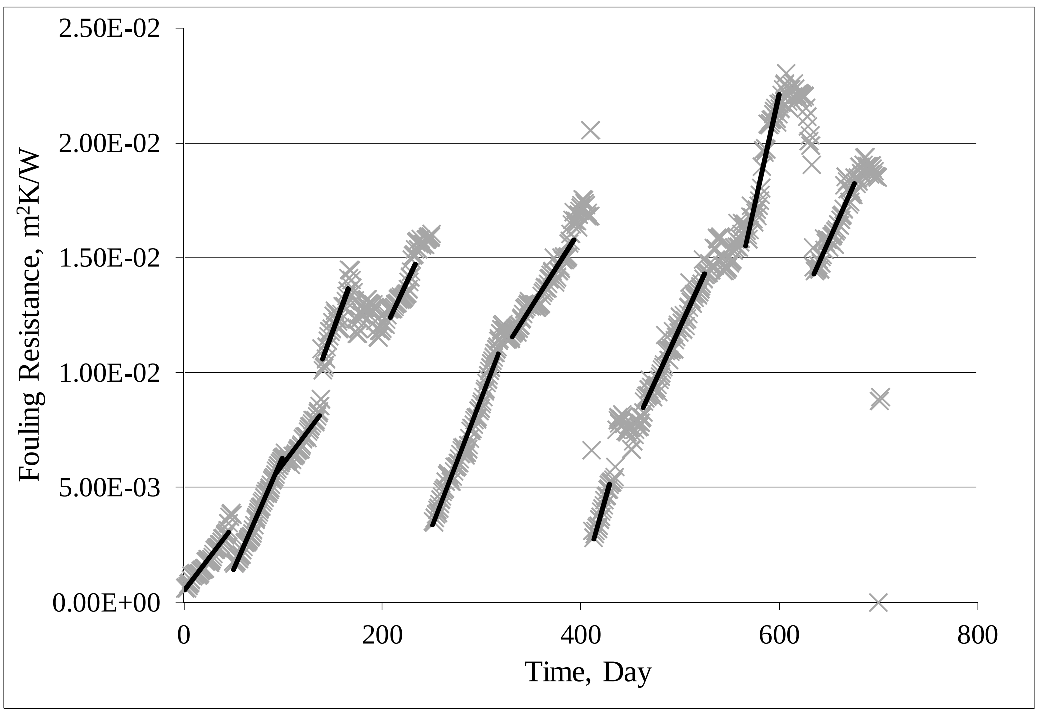

Figure 5.1: Observed fouling resistances for heat exchanger E-1A~D

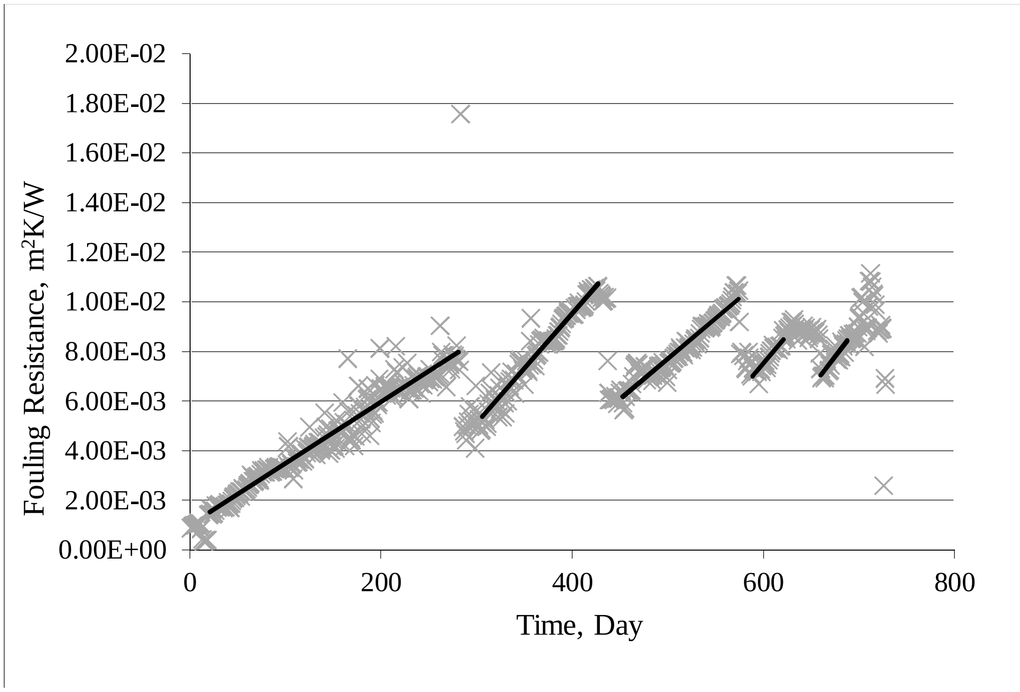

Figure 5.2: Observed fouling resistances for heat exchanger E-2.

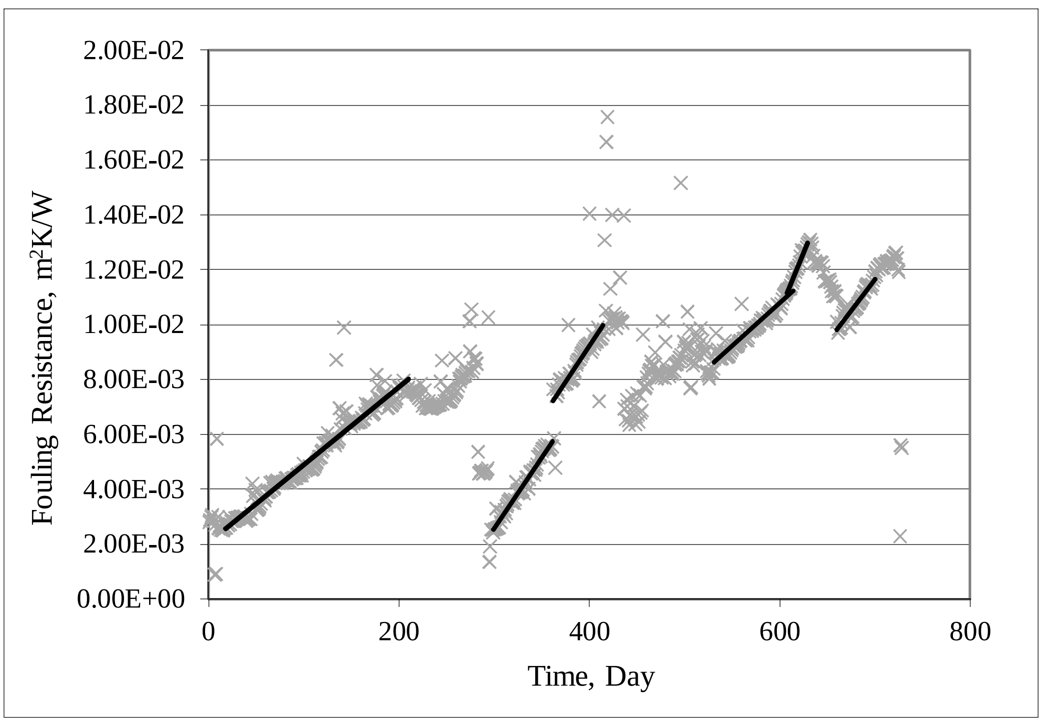

Figure 5.3: Observed fouling resistances for heat exchanger E-3.

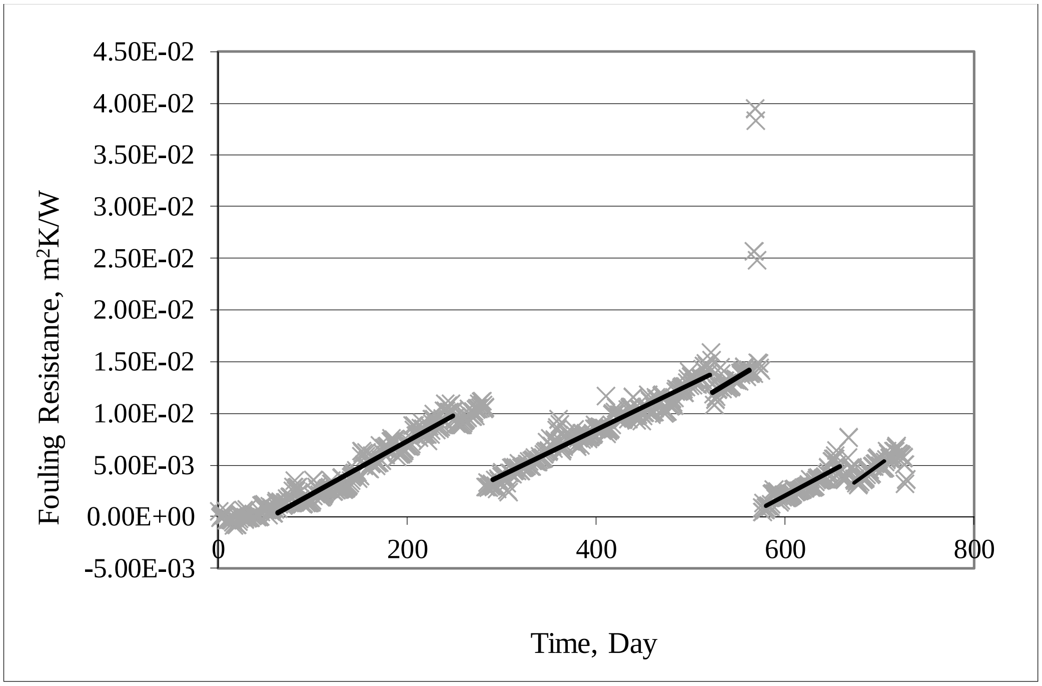

Figure 5.4: Observed fouling resistances for heat exchanger E-4A~D.

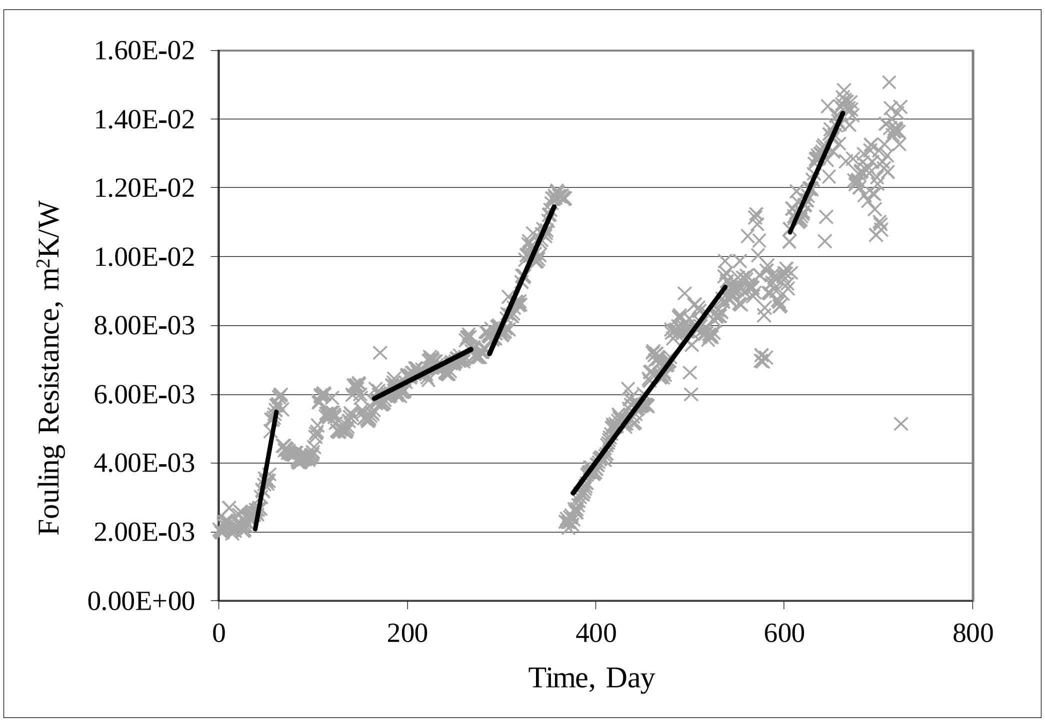

Figure 5.5: Observed fouling resistances for heat exchanger E-5.

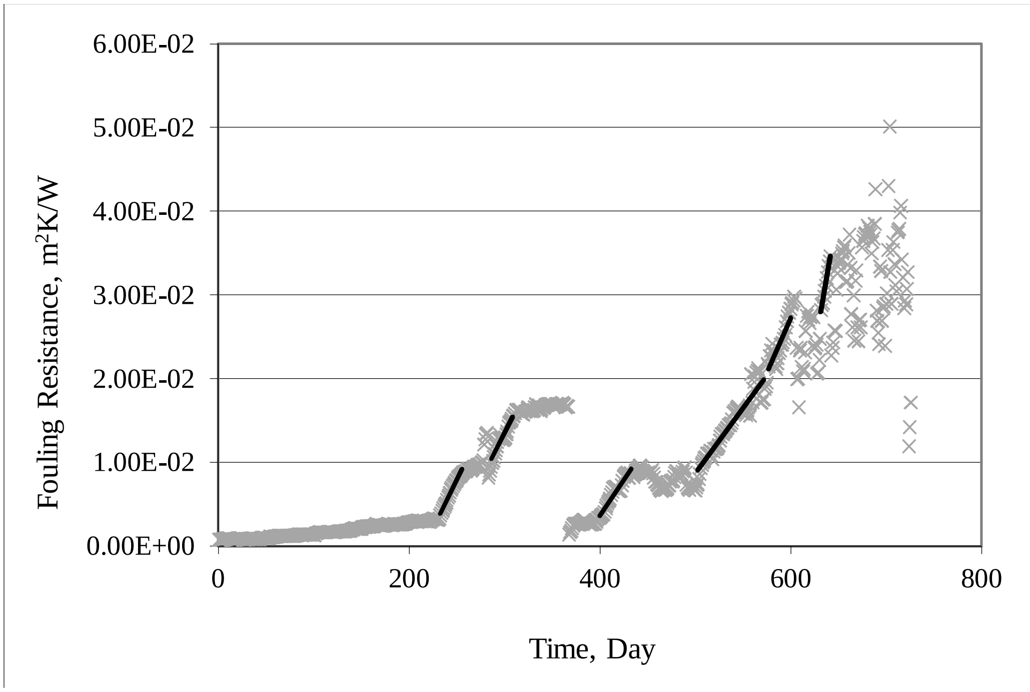

Figure 5.6: Observed fouling resistances for heat exchanger E-6.

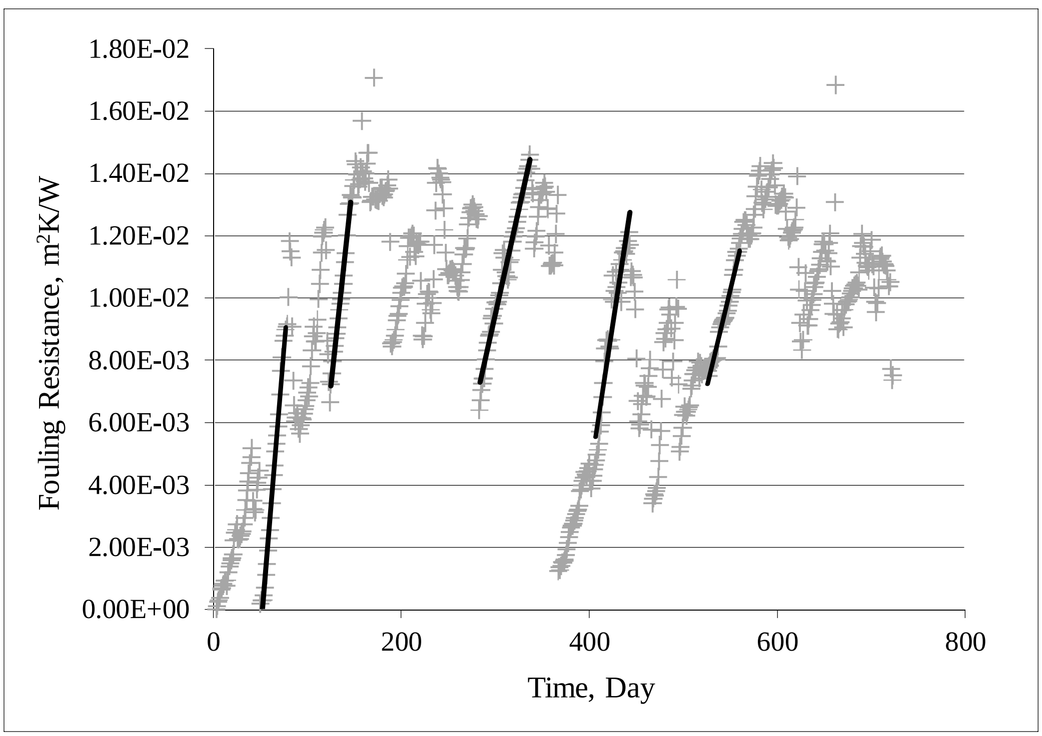

Figure 5.7: Observed fouling resistances for heat exchanger E-7A/B.

Table 5.3: Constant fouling rates for heat exchangers in the CPT

	Heat Exchanger	dR_f/dt (W/m^{2}K)/day
1.	E-1A~D	9.69E-05
2.	E-2	4.02E-05
3.	E-3	5.02E-05
4.	E-4A~D	5.26E-05
5.	E-5	6.05E-05
6.	E-6	2.10E-04
7.	E-7A/B	1.54E-04

The fouling rate normally increases with heat exchanger tube wall temperature and reduces with surface shear. Heat exchangers E-6, E-7 A/B, and E-1A~D have the highest fouling rates while E-5, E-4A~D and E-3 have relatively lower fouling rates. Heat exchanger E-2 has the lowest fouling rate throughout the operating period. During the optimisation study, the fouling resistances for the heat exchangers were assumed to be 0.000001 W/m^{2}K after the mechanical cleaning of the heat exchanger.

5.3 Results and Discussions

The cleaning schedule optimization for the CPT was performed for a period of two years. The inputs to the CPT model are varied daily based on the collected historical operating data from an operating refinery. These inputs include: i) operating temperatures at boundary conditions, and ii) flow rates of process and preheating fluids at the boundaries.

5.3.1 Base Case

A base case of cleaning schedule for the CPT was estimated based on the assumptions: i) the heat exchangers must be cleaned when the fouling resistance of the heat exchanger is more than the maximum allowable limit as specified in Table 5.2, and, ii) constant linear fouling rates are used to predict the fouling resistances for each heat exchanger (as discussed in section 5.2.1). As a result, the costs of HEN for base case are determined as shown in Table 5.4.

Table 5.4: Operating costs of heat exchangers in CPT for base case.

Description	Cost, mil RM/yr
1. Total un-recovered energy	5.711
2. Total cleaning costs	1.387
Total Operating cost	7.097

Table 5.5: Durations between cleanings for heat exchangers in CPT base case.

	Duration between cleanings, days						
Case	E-1A~D	E-2	E-3	E-4A~D	E-5	E-6	E-7A/B
Base Case	249	266	266	285	120	120	128

Table 5.6: Cleaning schedule of base case for 2 years of CPT operation.

	M 1	M 2	M 3	M 4	M 5	M 6	M 7	M 8	M 9	M 10	M 11	M 12
E-1A~D												
E-2												
E-3												
E-4A~D												
E-5												
E-6												
E7A~B												
	M 13	M 14	M 15	M 16	M 17	M 18	M 19	M 20	M 21	M 22	M 23	M 24
E-1A~D												
E-2												
E-3												
E-4A~D												
E-5												
E-6												
E7A~B												

*M = month

As observed from Table 5.4, total operating cost of HEN is RM 7.097 mil/yr, in which major portion of the cost is contributed by unrecovered energy cost, which is RM 5.711 mil/yr (80.47% of the total operating cost). Cleaning cost of heat exchangers is only RM 1.397 mil/yr (19.53% of the total operating cost).

The duration between cleanings for heat exchangers and estimated cleaning schedule of base case are shown is Tables 5.5 and 5.6. As observed in Table 5.6, the cleaning cycle for each heat exchanger under base case is not distributed evenly as each of the heat exchanger has different maximum allowable fouling resistance limit except for heat exchangers that shared the same bypass facility. As observed in Table 5.6, E-5 and E-6 need to be cleaned 3 times yearly, followed by E-7A/B that needs to be cleaned 2 times in the first year and 3 times in second year of CPT operation. The remaining heat exchangers need to be cleaned 1 time every operation year at different intervals of duration.

5.3.2 Optimization Cases

The optimum cleaning schedule of CPT was obtained by minimizing the total operating costs based on the objective function as discussed in Section 5.2.2. The duration between cleanings is decreased based on the number of cleanings. In this optimization problem, the duration between cleanings are fixed as in Table 5.7. This assumption may be restrictive, but as a first attempt to solve the optimization problem with minimum computational load, it is considered a right choice.

Table 5.7: Number of cleanings in a period of 2 years and the corresponding fixed duration between cleanings

Duration between cleanings, days	Number of Cleanings
365	2
243	3
183	4
146	5
122	6

There are totally 4 different heat exchanger combinations of heat exchangers that needed to be cleaned. These heat exchanger combinations are E-1A~D, E-2 & E-3, E-4A~D, and E-7A~D. Based on the opinion from engineers of the refinery, the minimum duration between cleaning must at least 120 days. Therefore, duration of cleanings of E-5 & E-6 that shared the same bypass facility are not manipulated in this optimization study as its duration of cleanings reach the minimum duration between cleaning.

During the initial iteration, the duration between cleanings of the heat exchanger combinations are maintained at the base case. Duration between cleaning more than the base case will not be considered with the assumption that the hydraulic flow limitations will occur at maximum fouling resistance constraint.

Table 5.8 shows the results of optimization. 24 runs were performed in 6 iterations to determine the optimal cleaning schedule corresponding to the minimum operating cost of CPT.

Table 5.8: Iteration results of optimization problem.

Iterations		Duration between cleanings, days E-1 A~D	E-2	E-3	E-4 A~D	E-5	E-6	E-7A/B	Additional Energy, RM/2 years	Cleaning Cost, RM/2 years	Total, RM/2 years
Iteration 1	**Run1**	**183**	266	266	285	120	120	128	**10,536,671**	**3,708,026**	**14,244,696**
	Run2	249	**243**	**243**	285	120	120	128	10,583,302	3,828,184	14,411,486
	Run3	249	266	266	**243**	120	120	128	10,830,765	3,503,655	14,334,420
	Run4	249	266	266	285	120	120	**122**	10,773,340	3,777,604	14,550,944
Iteration 2	Run5	**146**	266	266	285	120	120	128	10,365,875	3,984,242	14,350,117
	Run6	183	**243**	**243**	285	120	120	128	**10,523,175**	**3,737,756**	**14,260,931**
	Run7	183	266	266	**243**	120	120	128	10,591,908	3,807,920	14,399,828
	Run8	183	266	266	285	120	120	**122**	10,515,187	3,777,604	14,292,791
Iteration 3	Run9	**146**	243	243	285	120	120	128	10,308,383	4,013,973	14,322,356
	Run10	183	**183**	**183**	285	120	120	128	10,410,052	3,839,145	14,249,198
	Run11	183	243	243	**243**	120	120	128	**10,366,291**	**3,837,651**	**14,203,942**
	Run12	183	243	243	285	120	120	**122**	10,492,638	3,737,756	14,230,394
Iteration 4	Run13	**146**	243	243	243	120	120	128	10,249,677	4,113,867	14,363,544
	Run14	183	**183**	**183**	243	120	120	128	**10,256,008**	**3,939,040**	**14,195,048**
	Run15	183	243	243	**183**	120	120	128	10,216,664	4,048,540	14,265,204
	Run16	183	243	243	243	120	120	**122**	10,351,451	3,907,229	14,258,680
Iteration 5	Run17	**146**	183	183	243	120	120	128	10,086,829	4,215,256	14,302,086
	Run18*	**183**	**146**	**146**	**243**	**120**	**120**	**128**	**9,918,786**	**4,020,151**	**13,938,937**
	Run19	183	183	183	**183**	120	120	128	10,128,994	4,149,929	14,278,922
	Run20	183	183	183	243	120	120	**122**	10,195,303	3,907,229	14,102,532
Iteration 6	Run21	**146**	146	146	243	120	120	128	10,018,850	4,296,367	14,315,217
	Run22	183	**122**	**122**	243	120	120	128	**10,048,152**	**4,141,818**	**14,189,970**
	Run23	183	146	146	**183**	120	120	128	10,010,616	4,231,040	14,241,656
	Run24	183	146	146	243	120	120	**122**	10,147,978	4,089,729	14,237,707

**Optimized case*

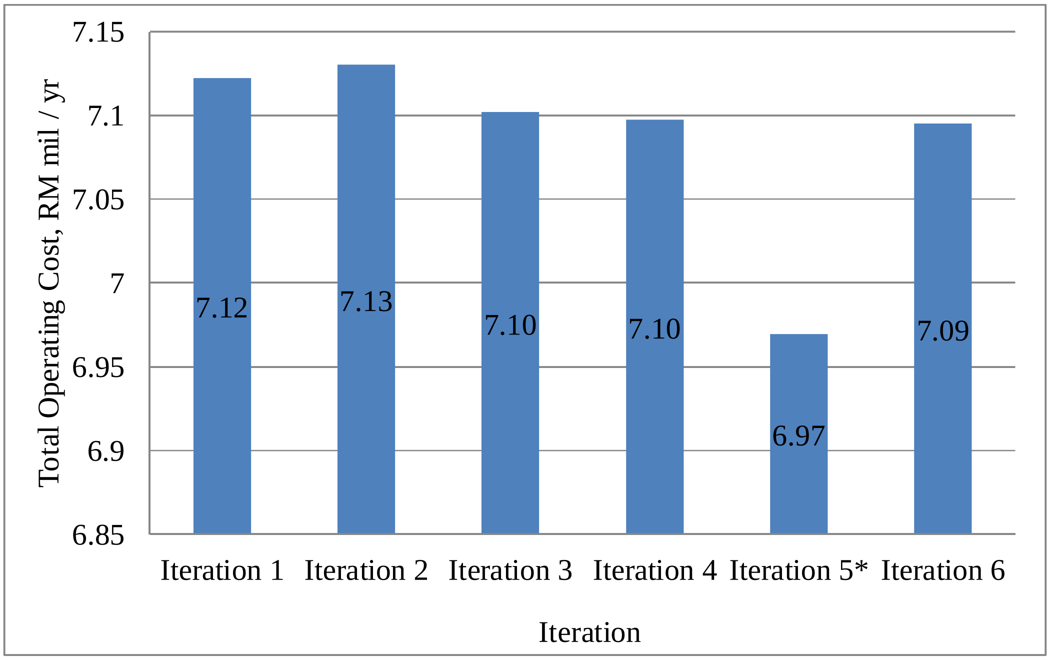

Figure 5.8: Results of Optimization Iterations

Fig. 5.8 showed the results of optimization for the 6 iterations. As compared to the operating cost of base case, the optimum cleaning schedule reduced the operating cost from RM 7.097 mil to RM 6.97 mil per year with a total operating cost saving of RM127, 682 per year.

Table 5.9 indicates that the optimal cleaning schedule for heat exchangers in CPT for 2 years of total simulation time.

Table 5.9: Optimized cleaning schedule for heat exchangers in CPT in 2 years total simulation time

	M 1	M 2	M 3	M 4	M 5	M 6	M 7	M 8	M 9	M 10	M 11	M 12
E-1A~D												
E-2												
E-3												
E-4A~D												
E-5												
E-6												
E7A~B												
	M 13	M 14	M 15	M 16	M 17	M 18	M 19	M 20	M 21	M 22	M 23	M 24
E-1A~D												
E-2												
E-3												
E-4A~D												
E-5												
E-6												
E7A~B												

The result indicates that it is not necessary to clean all the heat exchangers with higher fouling rates more frequently. Optimization study showed that E-5 and E-6 should be cleaned 3 times per year, while E-7A/B should be cleaned 2 times in the first year and 3 times in the second year. The numbers of cleanings for these last three heat exchangers in the CPT are matching against the fouling rates of these heat exchangers, which E-6 has the highest fouling rates, followed by E-7A/B. As observed from Table 5.9, E-2 and E-3 needs to be cleaned twice yearly. E-1A~D should be cleaned twice a year. E-4A~D has the least cleaning requirement, which is once yearly. As discussed in Chapter 4, log-mean temperature differences (*LMTD*) play an important role in arresting the decrease or even improving the heat transfer performances fouling resistance increases in the heat exchangers in CPT will e. Besides, position of heat exchanger in CPT also influences the cleaning requirement of heat exchangers. Hence, the number of cleanings for E-1A~D, E-2, E-3 and E-4A~D are the balanced effects of fouling resistance, *LMTD* effects, position of heat exchangers and the sharing of bypass facilities between heat exchangers. The optimization yields a better total operating cost of CPT if E-1A~D, E-2, E3 and E-4A~D are cleaned before their fouling resistance reaches maximum allowable fouling resistance. This observation indicates that the normal operational ad-hoc decision to clean the heat exchanger when it reaches its pressure drop limitations might not economical for heat exchangers in CPT.

5.4 Summary

An optimization problem for the cleaning schedule of CPT was formulated and solved for a period of two years. Minimum operating cost was achieved considering the total unrecovered energy cost and cleaning cost of the heat exchangers. The optimization of cleaning schedule of heat exchangers is able to achieve a total cost saving of RM 127, 682 per year. The results of optimization problem indicate that the key factors to determine the optimum cleaning cycles are fouling rates and heat transfer interactions between heat exchangers for each heat exchangers in CPT. As a norm, heat exchangers with higher fouling rates should be cleaned more frequently to ensure optimum heat transfer efficiency across the CPT. Nevertheless, the results indicate that it is necessary for some of the heat exchangers in CPT to be cleaned before it reaches maximum allowable fouling resistance.

CHAPTER 6

CONCLUSION AND RECOMMENDATIONS

6.1 Conclusions

A crude preheat train (CPT) was selected in this research as a case study for analysing the effect of fouling and to determine an optimal cleaning schedule. Design and operational data for a period of two years was collected from a refinery. Based on the information collected, a rigorous CPT model was successfully developed under Petro-SIM™ environment. In addition, the fouling models were also developed to describe the fouling in the heat exchangers.

A simulation study was performed to investigate the effects of fouling on the thermal performance of individual heat exchangers in the CPT and that of overall CPT. The fouling resistances of the heat exchangers were estimated using semi-empirical fouling models. Same operating conditions were input to inlet streams of the CPT model. Heat transfer performances of heat exchangers were observed for a period of 300 days of simulation. The heat transfer performance dropped for some of the heat exchangers, and increased for other heat exchangers even though the fouling was becoming more severe. The major observations/conclusions based on the simulation study are listed below.

- *LMTD* effects are observed in heat exchangers of the CPT. The increase and decrease in the inlet temperatures for hot fluids and the process fluid, respectively, to a heat exchanger due to the lower heat transfer performance in the fouled upstream heat exchanger increases the temperature driving forces and hence improves the heat transfer performance of heat exchanger in CPT.
- Fouling of first heat exchanger in CPT contributes to the highest drop of heat

transfer performance of overall CPT.

Under the optimization study, an optimization problem was formulated and solved for a period of two years to determine the cleaning schedule of CPT that have the total minimum operating cost. Constant linear fouling rates were assumed for the heat exchangers in CPT. A base case was developed to investigate the basic cleaning schedule required to maintain the operation of CPT. The observations/conclusions of the optimization study are below.

- A savings of RM 127, 682 per year in operating cost.
- Fouling rates, heat transfer interaction between heat exchangers, and duration between cleanings are key factors in determining the cleaning schedule of heat exchanges in CPT.
- For certain instances, it is necessary for some of the heat exchangers to be cleaned before it reaches maximum allowable fouling resistance corresponding to maximum allowable pressure drop across the heat exchanger to achieve a cleaning schedule of CPT that yields minimum total operating cost.

6.2 Recommendations

Mixed integer programming approach was used to solve the optimization problem based on rigorous simulation model of CPT. As the simulation of the rigorous CPT model is computationally intensive and the number of function evaluations is quite large, more powerful optimization algorithms like genetic algorithm can be used to solve the optimization problem to identify the global minimum with less computational load. The optimization variables can also be changed from the number of cleanings in a fixed period to the duration between the cleanings which will be more appropriate for continuous operating plants.

During the periods when a heat exchanger is taken out of service for cleaning, the HEN configuration changes significantly and the balance of heat integration changes.

A systematic study to analyse the effect of changes in HEN configuration is recommended.

The effects of hydraulic limitations are not studied thoroughly in this study. A dynamic study can be performed to investigate the effect of hydraulic to HEN model under fouling condition.

REFERENCES

[1] ESDU (Engineering Sciences Data Unit), "*Heat Exchanger Fouling in the Pre-heat Train of a Crude Oil Distillation Unit*," in ESDU Intl, USA, 2000, Data Item 00016.

[2] H. S. Muller, M. R. Malayeri and A. P. Watkinson, "*Heat Exchanger Fouling; Environment Impacts*," in Heat Transfer Engineering, UK, 2009, 773-776.

[3] K. J. Bell, and A. C. Mueller, "*Wolverine Engineering Data Book II*," in Wolverine Tube Inc, UK, 1984, 31-38.

[4] H.M. Steinhagen, "*Heat Exchanger Fouling: Mitigation and Cleaning Technologies*," in Publico Publications, Germany, 2000, 24-25.

[5] J.M. Chenoweth, "*Final report of the HTRI/TEMA joint committee to review the fouling section of the TEMA standards*," in Heat Transfer Engineering, USA, 1990, 73–107.

[6] B. Peng, Q. W. Wang, and C. Chang, "*An Experimental Study of Shell and Tube Heat Exchanger With Continuous Helical Baffles*," in AEME, Xi'an Jiaotong University, China, 2007.

[7] A. K. Sheikh, S.M. Zubair, M. Younas, and M.O. Boudair, "*A Rish Based Heat Exchanger Analysis Subject to Fouling Part II: Economics of Heat Exchangers Cleaning*," in Elsevier Science Ltd, Saudi Arabia, 2000, 445-461.

[8] D. J. Kukulka, R. E. Baier, and J. C. Mollendorf, "*Factors Associated with Fouling in Process Industry*," in Heat Transfer Engineering, USA, 2004, 23–29.

[9] B. D. Crittenden, S. T. Kolaczkowski, and T. Takemoto, "*Use of tube Inserts to Reduce Fouling From Crude Oils*," in AIChE Symp. Series, Atlanta, 1993, 295, 89, 300- 307.

[10] M. Markowski, "*Reconstruction of a Heat Exchanger Network under Industrial Constraints -The Case of a Crude Distillation,*" in Elsevier Science Ltd, Applied Thermal Engineering, Poland, 2000, 20:1535-1544.

[11] C. Rodriguez, and R. Smith, "*Optimization of Operating Conditions for Mitigating Fouling in Heat Exchanger Networks,*" in Chemical Engineering Research and Design Trans IChemE, 2007, UK, Part A.

[12] M. H. Panjeshahi, and N. Tahouni, "*Pressure Drop Optimization in Debottlenecking of Heat Exchanger Networks,*" in Elsevier, Science Ltd, Energy, Iran, 2008, 942-951.

[13] B. L. Yeap, "*Design of Heat Exchanger Networks with Fouling Mitigation*," in CGPS dissertation, University of Cambridge, UK, 2001, 28.

[14] D. Bouris, E. Konstantinidis, S. Balabani, D. Castiglia and G. Bergeles, "*Design of a novel, intensified heat exchanger for Reduced Fouling Rates*," in International Journal of Heat and Mass Transfer 48, Greece, 2005, 3817–3832.

[15] W. A. Ebert, and C. B. Panchal, "*Analysis of Exxon crude-oil slip stream coking data*," in Fouling Mitigation of Industrial Heat-Exchange Equipment, Begell House, 1995, USA, 451–460.

[16] K. Brodowicz, M. Markowski, "Calculation of Heat Exchanger Networks for Limiting Fouling Effects in the Petrochemical Industry," in Applied Thermal Engineering, Elsevier Science Ltd, Poland, 2003, 23:2241–2253.

[17] T.R. Bott, "*Fouling of Heat Exchanger*," in Elsevier Science B.V., USA, 2005, 7-32.

[18] F. Smaīli, D. K. Angadi, C. M. Hatch, O. Herbert, V. S. Vassiliadis, and D. I. Wilson, "*Optimization of Scheduling of Cleaning in Heat Exchanger Networks Subject to Fouling: Sugar Industry Study Case*," in Trans ICheme, UK, 1999, Vol. 77, Part C.

[19] M. C. Georgiadis, and L. G. Papageorgiou, "*Optimal Energy and Cleaning Management in Heat Exchanger Networks Under Fouling*," in Trans Icheme, Greece, 2000, Vol 78-Part A.

[20] M. C. Georgiadis, and L. G. Papageorgiou, "*Optimal Cleaning of Heat Integrated Multipurpose Plants under Fouling Conditions*," in Elsevier, Science Ltd, Applied Thermal Engineering, Greece, 2001, 21:1675-1697.

[21] M. Markowski, and K. Urbaniec, "*Optimal Cleaning Schedule for Heat Exchangers in a Heat Exchanger Network*," in Applied Thermal Engineering, Poland, 2005, 25,1019-1132.

[22] S. Sanaye, and B. Niroomand, "*Simulation of Heat Exchanger Network (HEN) and Planning the Optimum Cleaning Schedule*," in Energy Conversion and Management, Iran, 2007, No. 48,1450-1461

[23] E. M. Ishiyama, W. R. Paterson, and D. I. Wilson, "*Thermal Effect of Fouling on Heat Transfer, Pressure Drop, and Throughput in Refinery Preheat Train: Optimization of Cleaning Schedules*," in Heat Transfer Engineering, UK, 2009, 805-814.

[24] Incropera, DeWitt, Bergman and Lavine, "*Fundamentals of Heat and Mass Transfer*," 6th Edition, Australia, 2002, Chapter 8.

[25] A. L. London, and R. A. Seban, "*A Generalization of the Methods of Heat Exchanger Analysis*," in Mechanical Engineering Department, Stanford University, Stanford, 1942,.

[26] D. M. Smith, "*Mean Temperature Difference in Crossflow*," in Engineering, UK, 1934, 479-481; 138, 606-607

[27] A. C. Mueller, "*Handbook of Heat Transfer*," in McGraw-Hill, New York, 1973, 35-80.

[28] O. Arsenyeva, L. Tovazhnyansky, P. Kapustenko, and G. Khavin, "*Mathematical Modeling and Optimal Design of Plate and Frame Heat Exchangers*," in Chemical Engineering Transactions, Ukraine, 2009, volume 18.

[29] M. A. S. Jeronimo, L. F. Melo, A. S. Braga, P. J. B. F. Ferreira, and C. Martins, "*Monitoring the Thermal efficiency of Fouled Heat Exchangers: A Simplified Method*," in Elservier, Science Ltd, Experimental Thermal and Fluid Science, Portugal, 1997, 14:455-463.

[30] B. D. Critenden, Kolaczkowski, and I. L. Downey, "*Fouling of Crude Preheat Exchangers*," in Trans. IChemE, UK, 1992, Vol 79: Part A, pp. 547-557.

[31] M. Srinivasan and A. P. Watkinson, "F*ouling of some Canadian Crude Oils*," in Heat Transfer Eng., Canada, 2005, vol. 26(1), pp. 7-14.

[32] D. Q. Kern, and R. E. Seaton, "*A Theoretical Analysis of Thermal Surface Fouling*," in Br Chem Eng, Germany, 1959, 258-262

[33] J. Aminian, and S. Shahhosseini, "*Evaluation of ANN Modeling for Prediction of Crude oil Fouling Behavior*," in Applied Thermal Engineering, Elsevier Science Ltd, Iran, 2008, No. 28- 668-674.

[34] V. R. Radhakrishnan, M. Ramasamy, H. Zabiri, V. D. Thanh, N. M. Tahir, H. Mukhtar, M. R. Hamdi, and N. Ramli, "*Heat Exchanger Fouling Model and Preventive Maintenance Scheduling Tool*," in Applied Thermal Engineering, Malaysia, 2007, No. 27-2791-2802.

[35] C. B. Panchal, W. C. Kuru, C. F. Liao, W. A. Ebert, and J. W. Palen, "*Threshold Conditions for Crude Oil Fouling, Proc. Understanding Heat Exchanger Fouling and Its Mitigation*," in Engineering Foundation Conf., Italy, 1997.

[36] G. T. Polley, D. I. Wilson, B.L. Yeap and S. J. Pugh, "*Evaluation of Laboratory Crude Oil Threshold Fouling Data for Application to Refinery Pre-heat Trains*," in Applied Thermal Engineering 22, UK, 2002, 777-788.

[37] N. Epstein, "*Optimum Evaporator Cycle with Scale Formation*," in Canadian J. Chem, Canada, 1979, 57:659-661.

[38] E. Casado, "*Model Optimizes Exchanger Cleaning," in Hydrocarbon Processing*," Oklahoma, 1990, 69(8), 71-76.

[39] A. K. Sheikh, S. M. Zubari, M. U. Haq, and M. O. Budair, "*Reliability Based Maintenance Strategies for Heat Exchangers Subject to Fouling*," in Trans ASME, Saudi Arabia, 1996, 118:306-312.

[40] E. Kothabasakis and B. Linhoff, "*Sensitivity Tables for the Design of Flexible Processes – (1) How Much Contingency in Heat Exchanger Design Networks is Cost Effective*," in Trans ICheme, Chem. Eng. Res Des, 1986, 64(A)-197-211.

[41] J. P. Holman, "*Heat Transfer*," 9th Edition, McGraw-Hill, New York, 2002.

[42] T. Kuppan, "*Heat Exchanger Design Handbook*," 2nd Edition, Taylor & Francis Group, New York, 2000.

[43] R. M. Pratt, "*Thermodynamic Properties Involving Derivatives*," in ASEE Chemical Engineering Division Newsletter, Malaysia, 2001.

[44] G. L. Kaes, "*Refinery Process Modeling: A Practical Guide to Steady State Modeling of Petroleum Processes*," The Athens Printing Company, Kaes Enterprises Inc., USA, 2000.

[45] KBC, "*Petrofine Physical Properties Manual*," KBC Advance Technology plc., 1999.

[46] M. R. J. Nasr, and M. M. Givi, "*Modeling of Crude Oil Fouling in Preheat Exchangers of Refinery Distillation Units*," in Elsevier Science Ltd, Applied Thermal Engineering, Iran, 2006, 26:1572–1577.

[47] D. I. Wilson, G. T. Polley, and S. J. Pugh, "*Mitigation of Crude Oil Preheat Train Fouling by Design, Heat Transfer Engineering*," in Taylor & Francis, UK, 2002, 23:24–37.

[48] S. Asomaning, C. B. Panchal and C. F. Liao, "*Correlating Field and Laboratory Data for Crude Oil Fouling*," in Heat Transfer Engineering, 2000, Taylor & Francis, USA, 21:17-23.

[49] B. D. Crittenden, Kolaczkowski, and I. L. Downey, "*Fouling of Crude Oil Preheat Exchanegrs*," in Trans. IChemE, UK, 1992, Vol 70, Part A- 547-557.

PUBLICATION

1. W.C. Low, M. Ramasamy. "An Analysis of Effect of Fouling on Heat Exchanger Network Performance", 24th Symposium Of Malaysian Chemical Engineers / 1st Int. Conf. on Process and Advance Materials – 1st Int. Conf on Process Eng. And Advance Materials, Kuala Lumpur, 2010.