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ANALYSIS OF FUEL REDUCTION STRATEGIES FOR CRUDE DISTILLATION UNIT

by

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ANALYSIS OF FUEL REDUCTION STRATEGIES FOR CRUDE DISTILLATION UNIT

by

NUR IZYAN BINTI ZULKAFLI

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PERAK

FEBRUARY 2011

DECLARATION OF THESIS

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DEDICATION

This work is dedicated to
My beloved husband, Muhammad Helmi bin Abu
My parents, Zulkafli bin Ismail and Zainab binti Abdul Aziz
My parents in-law, Abu bin Yusup and Narimah binti Mahbob
My brothers and sisters

ABSTRACT

There is greater awareness today on the depleting fossil energy resources and the growing problem of atmospheric pollution. Engineers are developing practical techniques to ensure energy processes are designed and operated efficiently. Inefficient furnaces and heat exchangers contribute to the problem due to higher fuel demand and higher carbon emission. In crude preheat train (CPT), fouling causes the reduction of heat transfer efficiency, which leads to higher furnace fuel consumption, and exert additional cost for heat exchanger cleaning and lost production. This thesis presents strategies to reduce fuel consumption in the furnace, which will lead to reductions of operational cost and environmental emission. The method of exergy analysis is applied to determine the baseline energy efficiency of the furnace and CPT in a crude distillation unit (CDU). The strategies consist of locating and reducing exergy lost through process modifications of the energy system and developing optimum scheduling for retrofit and/or cleaning of heat exchangers in the CPT. There are two options for achieving fuel savings in the furnace. The options are reduction of heat lost from furnace stack and enhancement of heat recovery in the CPT. The second option involves plant shutdown for overall cleaning of CPT (Case 1), online cleaning of heat exchangers (Case 2) and combined online cleaning with retrofit of high efficiency heat exchangers (Case 3). Reduction of heat loss from furnace stack contributes to the smallest cost saving of 6.44% without carbon credit. With carbon credit, the saving is increased to 6.70%. The largest energy and carbon dioxide emission savings are found from Case 3. The installation of high efficiency heat exchangers improves furnace inlet temperature (FIT) from 215^{o}C to 227^{o}C. Furthermore, Case 3 results in the highest percentage of cost saving by about 71% and 62% with and without carbon credit, respectively. The payback period for investment in high efficiency heat exchangers is 3 months, with carbon credit, and 4 months, without carbon credit, respectively. Thus, Case 3 is the most cost effective option for reductions of energy consumption and carbon dioxide emission in the CDU.

ABSTRAK

Kesedaran yang lebih besar hari ini mengenai sumber tenaga fosil yang berkurangan dan masalah pencemaran udara yang bertambah membuatkan jurutera membangunkan teknik-teknik praktikal bagi memastikan proses tenaga dirancang dan dikendalikan secara cekap. Dandang dan alat penukar haba yang kurang cekap menyumbang kepada masalah permintaan bahan bakar dan pembebasan karbon yang lebih tinggi. Dalam rangkaian pemanasan minyak mentah (CPT), pengotoran penukar haba mengurangkan kecekapan perpindahan haba dan meningkatkan kos operasi disebabkan pembersihan alat penukar haba dan kehilangan pengeluaran loji. Tesis ini menganalisa strategi pengurangan penggunaan bahan bakar dalam dandang bagi mencapai pengurangan kos operasi dan pembebasan karbon ke alam sekitar. Kaedah analisis exergy diguna untuk menentukan tahap kecekapan tenaga di dandang dan CPT dalam CDU. Strateginya terdiri daripada mencari dan mengurangkan exergy yang hilang melalui proses pengubahsuaian sistem tenaga dan penjadualan yang optimum untuk penukaran dan/atau pembersihan alat penukar haba. Dua opsyen dicadangkan. Opsyen pertama adalah pengurangan haba yang hilang dari cerombong dandang dan pemulihan haba dari CPT. Opsyen kedua dibahagikan kepada pembersihan keseluruhan CPT (Kes 1), pembersihan alat penukar haba secara "on-line" (Kes 2) dan pembersihan "on-line" berserta pengantian kepada alat penukar haba kecekapan tinggi (Kes 3). Penjimatan tenaga terendah diperolehi dari strategi pengurangan haba dari cerombong dandang, iaitu penjimatan kos sebanyak 6.44% tanpa kredit karbon dan 6.70% dengan kredit karbon. Peratusan tertinggi penjimatan tenaga dan pembebasan karbon dioksida adalah Kes 3. Pemasangan alat penukar haba kecekapan tinggi dalam Kes 3 telah meningkatkan FIT dari 215^{o}C kepada 227^{o}C. Kes 3 juga adalah peratusan tertinggi dalam penjimatan kos sekitar 71% dengan kredit karbon dan 62% tanpa kredit karbon. Tempoh pengembalian modal bagi pelaburan penukar haba kecekapan tinggi adalah 3 bulan, dan tanpa kredit karbon adalah 4 bulan. Dengan demikian, Kes 3 adalah opsyen yang mempunyai kos yang paling efektif dari sudut penjimatan tenaga dan pembebasan karbon dioksida.

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NOMENCLATURES

A	heat transfer area of heat exchanger (m^2)
CER	certified emission reduction (RM/tonne)
C_{cl}	heat exchanger cleaning cost (RM/unit)
C_{fl}	furnace's fuel cost (RM/GJ)
C_{pl}	production lost cost due to plant shutdown (RM/unit)
C_{ph}	purchase cost for heat exchanger ($)
C_p	specific heat for fuel gas (kJ/kmol.^{o}C)
Cc	specific heat for cold stream (kJ/kmol.^{o}C)
Ch	specific heat for hot stream (kJ/kmol.^{o}C)
CO2	carbon dioxide emission (kg/h)
dRf	fouling rate (m2.^{o}C/kW)
Fc	cold stream flow rate (kg/h)
Fh	hot stream flow rate (kg/h)
FG	fuel gas flow rate (m^3/h)
FIT	furnace inlet temperature (^{o}C)
M	molar mass (kg/kmol)
m	mass flow rate (kg/h)
P	pressure (kPa)
R	ideal gas constant (kPa.m^3/kmol.K)
R_{CER}	revenue of certified emission reduction (RM/kg)
Rf	fouling resistance (m^2.^{o}C/kW)
dRf	fouling rates (m^2.^{o}C/kW)
dRfhex	fouling rates for high efficiency heat exchanger (m^2.^{o}C/kW)
T	temperature (K)
Th1	inlet hot stream temperature (^{o}C)
Th2	outlet hot stream temperature (^{o}C)
Tc1	inlet cold stream temperature (^{o}C)

Tc2	outlet cold stream temperature (^{o}C)
TFT	theoretical flame temperature (^{o}C)
TIT	tower inlet temperature (^{o}C)
Q	heat duty (kW)
U	overall heat transfer coefficient (kW.m^{2}/^{o}C)
Uc	overall clean heat transfer coefficient (kW.m^{2}/^{o}C)
Uf	overall fouled heat transfer coefficient (kW.m^{2}/^{o}C)
y^{cg}	binary variable for changing heat exchanger
y^{cl}	binary variable for cleaning heat exchanger
ΔEx_{source}	exergy source (kW)
ΔEx_{sink}	exergy sink (kW)
ΔEx_{lost}	exergy lost (kW)
ΔFG	extra fuel gas flow rate (m^{3}/h)
ΔH	enthalpy (kW)
ΔH_{c}	heat of combustion (kW)
ΔS	entropy (kW/^{o}C)

Subscript

i	heat exchanger
t	period

Greek letter

α	conversion factor (GJ.h/month.m^{3})
ρ	density (kg/m^{3})

CHAPTER 1

INTRODUCTION

1.1 Crude Distillation Unit

The crude distillation unit (CDU) is the first step in a refinery complex to separate crude oil into different fractions depending on the difference of boiling temperatures of the various constituents. In a typical CDU, the crude oil feed stream is preheated in a crude preheat train (CPT) in two sections. The first section runs from storage to a desalter unit. While, the second section runs from desalter to the furnace. The main function of the desalter unit is to remove salts, water and sediment present in the crude oil feed stream. CPT utilizes the high temperatures of the distillation column product streams. The crude outlet temperature of the CPT could reach up to 280^{o}C. Then, the crude oil is further heated up in the furnace. Typical furnace outlet temperature values are 350^{o}C to 380^{o}C. The heated crude oil enters crude distillation column. It is a long column that consists of many trays. These trays have bubbles or holes to allow vapour to pass through them. Crude fractions settle in the rectifying section trays and are drawn off at four liquid side cuts depending on their average boiling point. The side cuts are naphtha, kerosene, diesel and fuel oil. The liquid that is drawn in the bottom tray of distillation column is atmospheric residue. (Al Muslim, et al., 2003)

1.2 Fouling of Heat Exchangers

The accumulation of deposits on the surfaces of heat exchangers is known as fouling. In CPT, the deposits may be crystalline, particulate matter, chemical reaction and corrosive material. The formations of these deposits depend on the fluid passing through the heat exchanger.

Crystallisation fouling occurs due to deposition of dissolved salts which at process conditions becomes supersaturated at the heat transfer surface. Supersaturation may be caused by evaporation of solvent, cooling below solubility limit of solution, heating above solubility limit of solution and mixing of streams with different composition. Particulate fouling is the deposition of small suspended particles such as clay, silt or iron oxide on heat transfer surfaces. Fouling may also be caused by chemical reactions, where deposits are formed at the heat transfer surface. Meanwhile, the corrosion of the heat transfer surface will increase the surface roughness. The increased surface roughness of heat transfer area may promote corrosion fouling to occur (Steinhagen, 2000).

Figure 1.1 shows a shell and tube heat exchanger that is fouled with particulate matter. Figure (a) on the left is a badly fouled heat exchanger tubes while figure (b) on the right shows the tube condition after cleaning.

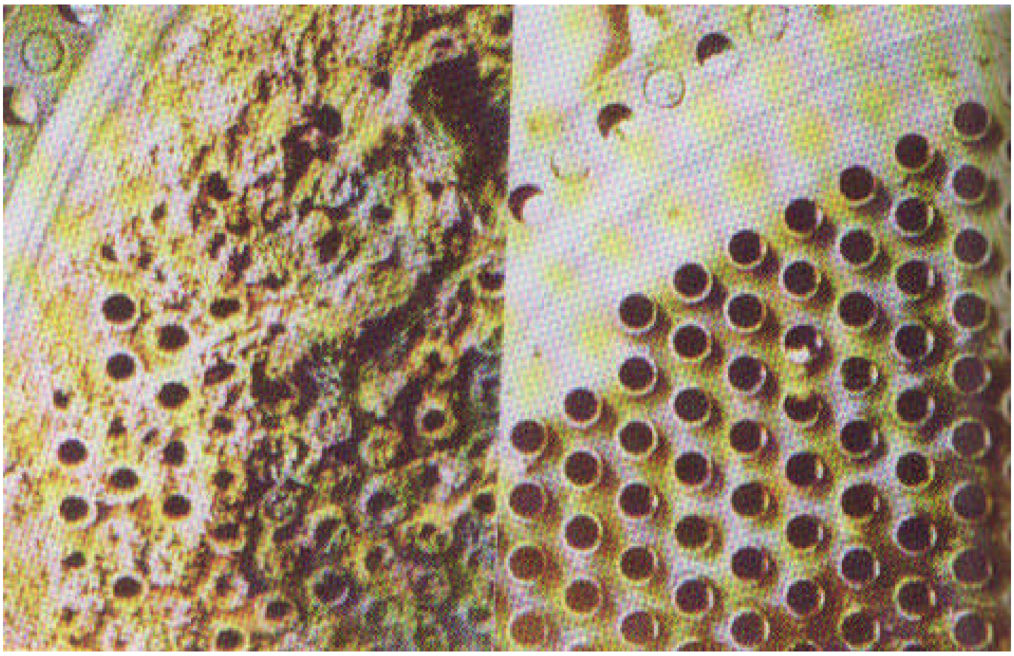

Figure 1.1 Picture of fouled (a) and clean (b) shell and tube heat exchanger

The progressing of fouling with time is ideally illustrated in Figure 1.2. Line A represents a linear relationship of deposit thickness with time. The linear relationship indicates that the rate of development of the fouling layer is constant along the period of time, *t*. Line B represents a falling rate of deposition once initiation has occurred. If the process of deposition are allowed to progress sufficiently an asymptotic curve would be produced as presented by line C (Bott, 1995).

The presence of deposits causes resistance to heat transfer and reduces the heat transfer efficiency of heat exchangers. The reduction of heat transfer efficiency in heat exchangers over time increases fuel consumption. The additional fuels are needed to overcome the shortfall in energy recovered due to fouling problem. The effects of fouling in heat exchangers not only increase the operational cost but also increase carbon dioxide (CO_2) that is emitted during fuel combustion. The increment of CO_2 brings adverse impact to the environment due to the effect of greenhouse gases emission that causes global warming on earth (Bott, 1995).

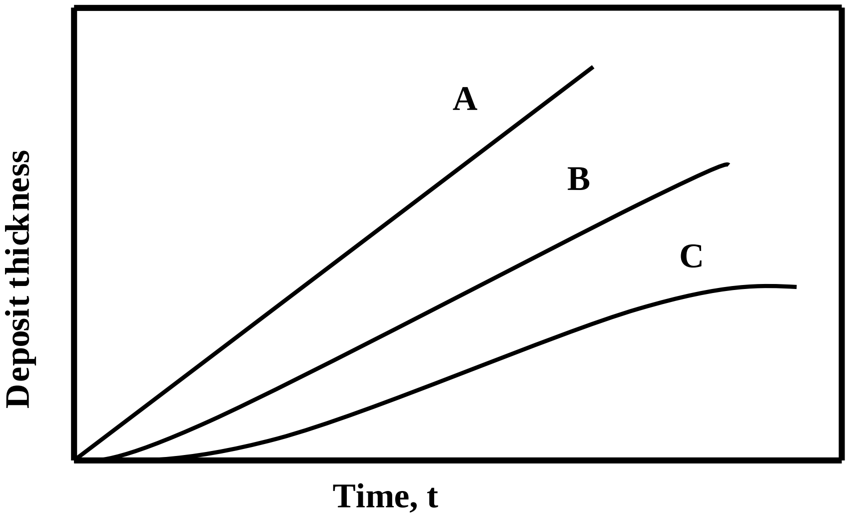

Figure 1.2 Idealised deposition curves

1.2.1 Heat Exchanger Cleaning

In order to maintain or restore heat transfer efficiency, it is necessary to clean the heat exchangers. The methods of cleaning are classified into two groups, namely online and offline cleaning. Online cleaning is performed during plant operation by isolating the heat exchangers through bypasses. Online cleaning can also be implemented for heat exchangers in parallel position where one heat exchanger is in operation mode while the other one is on standby mode. When the performance of heat exchanger in operation mode is reduced due to fouling, the online cleaning can be performed for fouled heat exchanger by putting the standby heat exchanger into operation mode. The fouled heat exchanger has an opportunity to be cleaned. At the same time, online cleaning allows for the production to be maintained. Meanwhile, offline cleaning is performed when the plant is shutdown during plan or unplanned shutdown (Bott, 1995).

The choice of cleaning techniques depends on the accessibility of the fouled surfaces and the fouling severity in the heat exchangers. The common mechanical cleaning technique in refinery is steam-blasting and hydro-blasting. If deposits are very tenacious, sand can be added to the pressurized water to increase the cleaning efficiency. On the other hand, using air or hydropressure, rubber plugs or metal scrapers can be shot through the tubes. Metal scrapers is shot through the tubes at water pressure of 35 bars and a scraper velocity of 3-6 m/s results in the removal of deposits. In general, water pressure systems are safer than air pressure system due to the compressibility and subsequent rapid expansion of gases (Bott, 1995).

Chemical cleaning may be required if the deposits are difficult to remove by mechanical cleaning. The selection of chemical and the cleaning procedure depend on the type of deposit and the configuration of the heat exchanger on economical and environmental concern. Most chemical cleaning consists of five distinctive stages. The five stages are alkaline cleaning, alkaline rinses, acid cleaning, acid rinses and passivation. The aim of alkaline cleaning is to remove the organic surface of the deposit in order to make the inorganic surface hydrophilic. Once the surface is hydrophilic, the deposit is softened and dissolved by application of the suitable acid blend. This blend usually contains inhibitor to prevent corrosion of the base metal by the acid. Before and after each chemical step, high flow water flushes are required to soften the deposit. The last stage is passivation where the protective oxide film is formed on the base metal to avoid corrosion (Steinhagen, 2000).

1.2.2 Selection of Heat Exchangers to Minimize Fouling

The major consideration for the selection of heat exchangers is the suitable techniques for cleaning heat exchangers. The method of cleaning should be a feature of the preliminary concept in designing the heat exchangers.

Table 1.1 lists some of the features of typical type of heat exchangers. It may be used as a preliminary guide in heat exchanger selection. For example, the shell and tubes heat exchanger is designed where the fluid that is likely to create higher fouling should be made to flow through the tubes sides of heat exchangers because tubes side is relatively easy to clean than the shell side. The fluid with lower fouling tendency should flow through the shell side of heat exchanger (Bott, 1995). The plate and frame heat exchanger may be attractive as it can be easily disassembled for cleaning and sterilising. The spiral heat exchanger performs well for fluids with a high concentration of suspended solids because low velocity region in this heat exchanger. Meanwhile, highly compact heat exchanger is normally avoided for severe fouling conditions as it is difficult to clean.

Table 1.1 Features of some typical types of heat exchanger

Type	Materials of construction	Ease of cleaning	Notes
Shell and tubes	Most materials	Tubes relatively easy to clean, shell more difficult	Widely used
Gasketted plate	Stainless steel	Easy to clean	Compact
Double pipe	Commonly in carbon steel	Inner tube relatively easy, annular space more difficult	Only useful for small heat transfer areas
Spiral	Most materials	Easy access to whole channel length	Compact: useful for slurries and fouling conditions
Plate-fin	Aluminium, stainless steel, titanium	Only chemical cleaning possible	Highly compact
Air-cooled	Commonly aluminium fins on carbon steel tubes	Inside tubes relatively easy, finned surface more difficult	Large plot area required

The most common heat exchanger applied in oil refinery is the shell and tube heat exchanger. The shell and tube heat exchanger is a reliable and well proven technology. It tolerates high design temperatures and pressures and can undertake almost any range of duty. However, the shell and tube requires a large heat transfer area as a result of a rather low overall heat transfer coefficient. In addition, the construction often involves the use of materials that are rather thick compared with plate type heat exchanger. The result is a large and heavy construction which entails high purchase and installation cost (Arvidsson, 2003).

One of the encouraging aspects for heat exchangers is the advances in development of compact heat exchanger. The use of compact heat exchanger in industry has its advantages in reducing energy consumption of operation and minimizing capital investment (Hesselgreaves, 2001).

One compact heat exchanger available in the market today is Alfa Laval's Compabloc welded plate heat exchanger. Figure 1.3 demonstrates the sectional view of Compabloc. The two media in the Compabloc heat exchanger flow in alternately welded channels between the corrugated plates. These corrugated plates promote high turbulence which provides high heat transfer efficiency and help minimize fouling. The corrugated plates are welded alternately to form channels. By simply unbolting the side panels, it is possible to gain access for inspection, service or cleaning by hydro blasting with a high pressure water jet. The media flows in a cross-flow arrangement within each pass as shown in Figure 1.3. Each pass is separated from the adjacent passes by a pressed baffle which forces the fluid to turn between the plate pack and the panel (Gunnarsson et al., 2008).

The high turbulence created by the corrugated pattern in the Compabloc results in very efficient heat transfer and minimises fouling problem. The heat transfer coefficient in a Compabloc is normally 2-4 times higher than in shell and tube heat exchanger. The compactness of the Compabloc means that the hold up volume is very low. The low hold up volume provides quick startup, easy control and fast response time. The high heat transfer efficiency in the Compabloc makes a temperature approach closer than shell and tube heat exchanger. (Arvidsson, 2003).

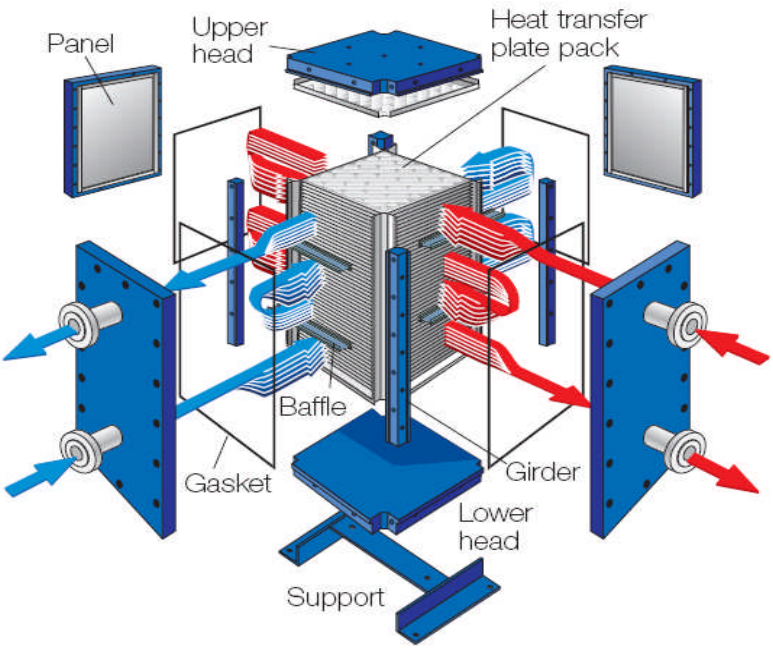

Figure 1.3 Sectional view of Compabloc

1.3 Furnace Operation

Oil refining and petrochemical plants involve extensive heating of hydrocarbon and other fluids. The fluid is heated by direct heating to achieve the required temperature. Direct heating is the process where the fluid under pressure that is contained in tubes is heated from the outside by direct exposure to flames. The oil refining and petrochemical furnaces commonly have two forms of furnaces. These are the cylindrical furnaces with one central burner or a ring of burners in the base and cabin furnaces with many burners in a row along the floor. The primary aim of a furnace operation is to attain a high processing temperature. In the furnace, heat is liberated by burning fuel with air or oxygen and some of this heat is transferred to the fluid. This process is known as combustion process where fuel and oxygen are burnt together at sufficiently high temperature to produce heat and combustion products. The combustion is completed when only carbon dioxide is produced. Complete combustion can be achieved by providing excess air. Incomplete combustion of fuel is associated with the formation of carbon monoxide. The products of combustion which leave the furnace through a furnace stack are known as flue gas. (Mullinger et al., 2008).

Figure 1.4 illustrates a schematic diagram of the crude furnace operation. The crude is preheated by crude preheat train before entering the furnace at furnace inlet temperature (FIT). The heat source is provided by the burning of fuel with air at theoretical flame temperature (TFT). TFT is a calculated temperature of flame. TFT is possible when the combustion is completely adiabatic. The heat from the burning of fuel with air is transferred to the crude oil. The heated crude oil enters the crude distillation column at tower inlet temperature (TIT). The remaining heat in the furnace leaves through the furnace stack at stack temperature (T stack).

The main contribution of high furnace efficiency is effective heat transfer and the heat transfer from flames is largely by radiation. The radiant heat transfer is proportional to the flame emissivity. The fuel with high flame emissivity is preferred for high temperature process because radiation normally contributes the greater proportion of heat transfer under this condition. When fuel with low flame emissivity is used, furnace must be designed to increase the contribution of convective heat transfer (Mullinger et al., 2008). Fuels with high carbon/hydrogen (C/H) ratios, for example oils and solid fuels, are more likely to burn with luminous flames due to high flame emissivity. Fuels with low C/H ratios, mostly gaseous fuels, tend to burn with clear flames due to low flame emissivity. The heat transfer from a luminous flame is usually greater than clear flame at the same temperature (Trinks, 2003).

Another contributing factor to furnace performance is the furnace atmosphere for optimum process efficiency. This is usually achieved by supplying excess air and generally requires very close control over fuel and air flow rates to ensure complete combustion. The preferred fuel for complete combustion in the furnace is gaseous fuels. There are processes that use solid fuel which contains excess coke and a carefully controlled air supply. However, carbon monoxide concentrations are very high typically above 20%. Another factor affecting furnace atmosphere is the sulphur content of the fuel. The products of combustion when excess air is present are SO_2 and SO_3. Generally SO_2 dominates but greater concentration of SO_3 can be formed when excess air is high. These gases are highly reactive as well as corrosive and can affect product quality by reacting with these gases (Mullinger et al., 2008).

Refractory is a major component of furnace. Furnace walls built of successive layers of hard refractory, insulating refractory and fibre block to reduce heat loss to the surroundings (Trinks, 2003). Refractory life may be also affected furnace performance. Fuel choice and furnace operating conditions can be the major influence on refractory life and hence on operating costs. The refractory failure often occurs for intermittent operation or cyclic operation. Furnace may start from cold at the beginning of each working week and than is heated during operation followed by a cooling phase after operation. The furnace is repeatedly heated and cooled which can lead to rapid refractory failure. However, this is not really a problem with a continuously operating furnace (Mullinger et al., 2008).

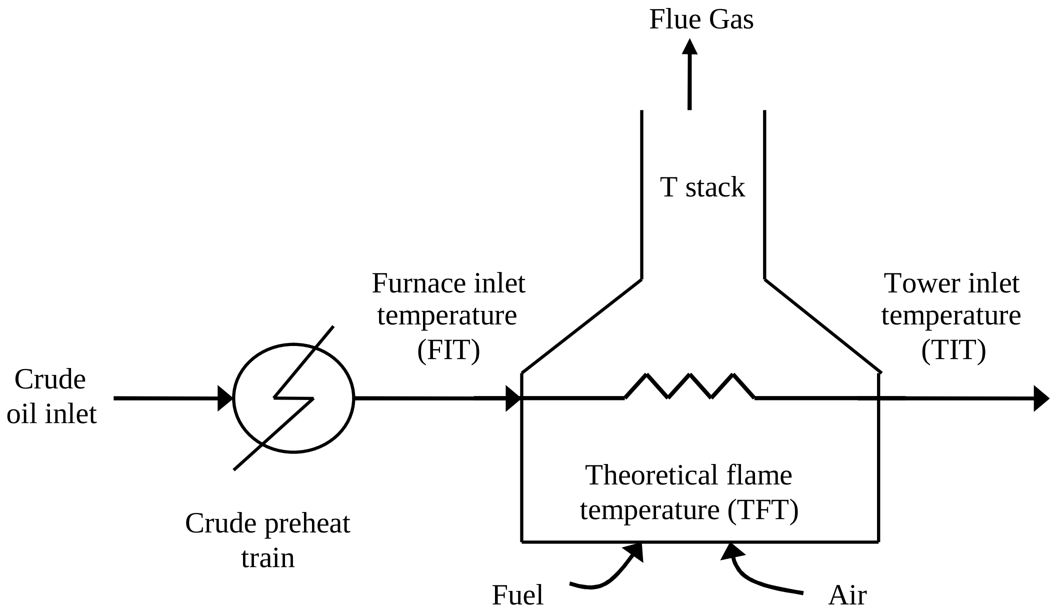

Figure 1.4 Schematic diagram of crude furnace operation

1.4 The Concept of Exergy

The exergy of the system is defined as a measure of the ability to do work or the work potential of a great variety of streams such as mass, heat, work that flow through a system at a specified reference environment. Exergy represents a quantitative measure of the usefulness or quality of energy. Exergy is conserved only when all processes in the system are reversible. Exergy is destroyed whenever an irreversible process occurs. Exergy is always destroyed partially or totally in real processes. The destroyed exergy is proportional to the generated entropy. Entropy is a measure of the amount of molecular disorder within a system. The amount of molecular disorder is characterized by the amount of disordered energy and its temperature. A system possessing a high degree of molecular disorder such as a high temperature gas has high entropy. The destroyed exergy is responsible for the less-than-ideal efficiencies of systems or processes. Exergy analysis is performed on a process to quantify the thermodynamic imperfections and process irreversibility (Dincer, et al., 2007).

It is important to differentiate between exergy and energy in order to avoid confusion with energy-based methods of thermal system analysis and design. Exergy depends not just on the state of a system or flow but also on the state of environment. The exergy of a system increases when it deviates from the environment. For example, a specified quantity of hot water has higher exergy content during winter than on a hot summer day. A system in complete equilibrium with its environment does not have any exergy content. According to the first law of thermodynamics, energy is neither lost nor destroyed but it is converted from one form to another. In the second law of thermodynamics, energy is degraded during the conversion which can be described as its ability to do work. The ability of energy to do work is represented as exergy (Dincer, et al., 2007).

The general form of exergy balance equation of a system is described as follows:

Total Exergy Input = Total Exergy Output + Total Exergy Loss

Figure 1.5 illustrates a Sankey diagram which represents the general exergy balance of a system. Sankey diagram is a special type of flow diagram with the size of the arrow showing the magnitude of the flow. In the figure below, the exergy transfers between processes is visualized between the exergy input and the exergy output through a system. Some of the exergy input is lost to the environment due to irreversibility of the processes.

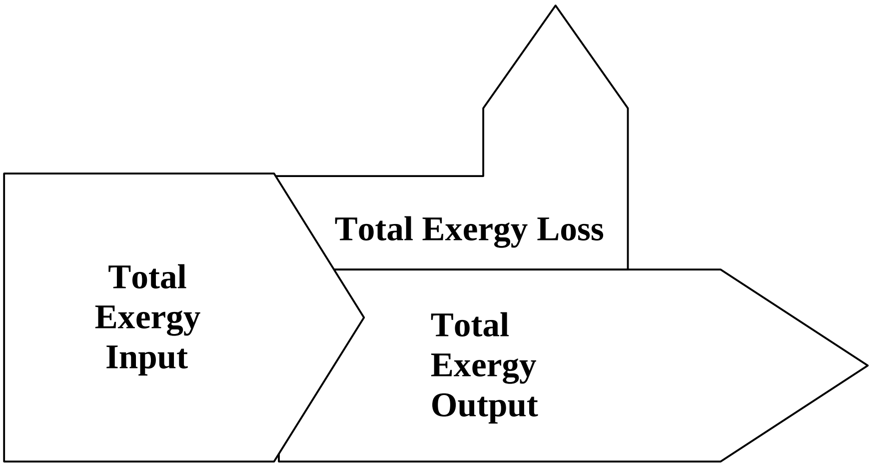

Figure 1.5 Sankey Diagram

Exergy analysis is an effective method to achieve efficient energy utilization with reduced environmental and sustainability impact. Exergy analysis has also been demonstrated for providing optimum design and operation. In this regard, exergy analysis appears to be an effective tool for addressing the impact of energy resource utilization on the environment, furthering the goal of more efficient energy resource utilization and enabling locations, types and true magnitudes of wastes and losses to be determined. In addition, exergy analysis can reveal whether or not and how much it is possible to design more efficient energy systems by reducing the inefficiencies in the existing systems (Dincer, et al., 2007).

1.5 Problem Statement

The accumulation of unwanted deposits on the surface of heat exchangers, or fouling, reduces the overall heat transfer efficiency in the CPT of a CDU. The reduction of heat transfer efficiency in CPT causes significant increment of fossil fuel consumption in the furnace. The higher consumption of fuel may be also caused by reduction of furnace efficiency due to greater heat loss to the surrounding. The high furnace stack temperature may contribute to greater furnace heat loss. The additional consumption of fossil fuel is needed to overcome the shortfall of energy recovered and this will affect the conservation of limited energy resources. The higher consumption of fossil fuel also causes environmental problem due to higher emission of carbon dioxide from combustion of fossil fuel to the atmosphere. Furthermore, the operational cost of the plant increases not only contributes to high consumption of fuel but also product yield and throughput reduction. In order to overcome this problem, heat exchangers are cleaned between shutdowns or during operations to restore its efficiencies. However, the most crucial factor for heat exchanger network cleaning schedule is to determine which heat exchanger need to be cleaned and when during operation. The loss of production that is caused by plant shutdown for cleaning is often more significant than the cost of heat exchangers cleaning. The time of which heat exchanger has to be cleaned is a scheduling problem where considerations between the cost of cleaning, the cost of fuel consumption and the cost of production lost need to be compromised.

1.6 Objectives

The main objective of this research is to develop fuel reduction strategies for CDU by exploring the modification of hot and cold stream temperatures of the furnace with respect to potential investment required. The second objective is to identify the most cost-effective option that results in the highest energy saving and carbon dioxide emission reduction.

1.7 Methodology

The relevant process data are extracted from a refinery to establish performance benchmark of the system. The base case data is used to conduct exergy analysis. From exergy analysis, the amount of exergy loss indicates the process inefficiency of the heat transfer equipments in CDU. The proposed options for fuel reduction strategies in this study are reduction of heat loss from furnace stack and enhancement of heat recovery in CPT. The enhancement of heat recovery in the CPT is divided into three cases of heat exchanger cleaning schedules. The cases are modelled and simulated to obtain optimal heat exchanger cleaning schedules, furnace inlet temperature (FIT) and extra fuel gas profiles. The options are proposed to minimize energy consumption and carbon dioxide emission as well as operational cost of the plant.

1.8 Scope of Study

This study is conducted on a typical refinery in Malaysia. The scope of this study is on the CPT and furnace in CDU. The relevant process data are extracted from the refinery to produce a set of base case data. Exergy analysis is conducted to establish the current performance of the system and to determine locations and magnitude of exergy losses in the CDU. A commercial software package, WORK software from the centre of process integration at the University of Manchester, United Kingdom is used to carry out exergy composite curve and exergy loss calculation. The models for optimum heat exchanger network cleaning schedules are coded in optimization software, General Algebraic Modelling System (GAMS) version 23.2 and solved by DICOPT solver. Mixed Integer Nonlinear Programming (MINLP) optimizer models are applied for all the cases. Some spreadsheet calculations are done by using Microsoft Office Excel 2003.

CHAPTER 2

LITERATURE REVIEW

This chapter presents literature review for the application of exergy analysis in industrial plant mainly in distillation processes. The previous work is a good basis for understanding the basic principles of exergy analysis and developing the improvement modification through the exergy analysis. This chapter also covers the previous work on the optimization of scheduling problems, the project on the reduction of carbon dioxide emission in industry and the application of Alfa Laval Compabloc heat exchanger in industry.

2.1 Application of Exergy Analysis in Industrial Plant

Many researches and engineers use exergy analysis as a method for analyzing, designing, improving and optimizing systems and processes. Some of the previous works done on the application of exergy analysis in industry has been reported on crude distillation unit, power plant, cogeneration plant, cooling system and fuel cell system.

Cornelissen (1997) used Aspen Plus to simulate the distillation process as well as the heat exchanger network in refinery crude distillation plant. The author showed that the exergy efficiency of the crude distillation unit was 27%, the vacuum distillation unit was 37.3%, heat exchanger network was 54.1% and the furnace was 54.1%. The author found that the major source of exergy loss was the usage of high quality fuel for producing middle quality heat in the furnace and the energy was wasted in atmosphere when cooling the products. The author proposed the usage of combined heat and power cycle to reduce the exergy loss by utilizing useful work loss in the distillation plant and anticipated that the reduction of exergy loss was 19% to 60%.

Doldersum (1998) also used exergy analysis to prove viability of distillation process modification. The author also used Aspen Plus program for process simulation and performed exergy analysis of a reaction and distillation section within a refinery. From the exergy analysis, the main part of the losses occurred in the furnaces and distillation columns. The author proposed several process modifications to reduce exergy loss which were decreasing operating pressure, using high pressure steam instead of furnace, splitting feed streams and recompressing overhead. With these process modifications total exergy losses reduced by 70% that directly resulted in a reduction of fuel of almost 40%.

Andersen et al. (2000) discussed about energy efficient distillation by optimal distribution of heating and cooling requirement. The author discussed that for a column having an infinite number of stages (N) with a heat exchanger on each stage would result in zero entropy generation in the distillation column. The entropy production reduced by 25% for installation of 2 extra heat exchangers, 37% for installation of 4 extra heat exchangers and 70% for installation of 19 extra heat exchangers. However, the authors did not provide any economic analysis for the proposed improvement option.

Dincer et al. (2002) utilized the use of energy and exergy analyses for coal-fired and nuclear electrical generating stations. From the result, steam generation sections of both stations were the highest exergy consumption. Energy efficiency and exergy efficiency of steam generation section were 95% and 49% for coal-fired generating station, and 95% and 42% for nuclear generating station. They emphasised that steam generation for both stations were significantly more efficient on an energy basis than on an exergy basis. Although 95% of the input energy was transferred to the preheated water, the energy was degraded as it was transferred. Exergy analysis highlighted this degradation.

Rivero (2002) described the past, present and future applications of exergy concept in petroleum refining and petrochemical industry. The application of exergy concepts allows to simultaneously considering the energy-economy-ecology model. The applications are grouped into two segments. The first is to integrate simulation, analysis and optimization program to conduct exergy analysis and to develop exergeoeconomic and exergeoecologic studies. The second segment refers to research and technological development of specific processes to consider energy, economy and ecology aspect where the justification of all the aspects is based on the exergy analysis.

Al-Muslim et al. (2003) conducted exergy analysis of single and two-stage crude oil distillation units. Al-Muslim used SimSci/PRO II program to simulate both single and two stage distillation units. This study showed that the overall exergy efficiency for single and two-stage distillation units were 14% and 31.5% respectively. The proposed two-stage distillation column showed 9.34% decrease in overall duty, 43.8% decrease in overall exergy losses and 125% increase in overall exergy efficiency compared to single stage distillation column. Nevertheless, no economic analysis is conducted for the proposed two stages crude distillation unit.

Another exergy analysis study by Rivero et al. (2004) on crude oil combined distillation unit and found exergoeconomic improvement potentials for all equipment of the process. The simulation of the plant was conducted using AspenPlus software and the exergy analysis subroutines developed by the IMP Exergy Group. The process streams of combined distillation unit consist of preheating and desalting section, topping section, stabilization section, atmospheric section and vacuum section. From the result, atmospheric section has the highest total exergy losses amounting to 60.54%. The reason for high exergy loss is due to its low effectiveness (24.48%). This section also has the highest improvement potential (68.89%). The highest cost of exergy loss is the atmospheric fired heater which is about 44.98% of the total cost of exergy loss. In their work, the authors did not propose any optimization approach to reduce the exergy loss.

In a study by Konoglu et al. (2004), exergy analysis was performed for open cycle desiccant cooling system. Relation between exergy destructions and exergy efficiency was proposed. That was, the smaller the exergy efficiency the greater the exergy destruction. From the result, desiccant wheel has the highest percentage of exergy destruction, followed by heating system, rotary regenerator and evaporative cooler. The authors proposed to replace evaporative cooler with reversible wet surface heat exchanger but no economic analysis was calculated.

Al Muslim et al. (2005) took one step ahead in their study on the effect of reference temperature on the exergy efficiencies and exergy losses in one and two-stage crude oil distillation unit. The results demonstrated that increasing reference temperature decreases the exergy efficiency in both one and two stage crude oil distillation systems and also increases the difference between the exergy efficiencies of both systems.

Hussian et al. (2005) performed a parametric study to examine the effect of varying operating conditions on the energy and exergy efficiency of polymer electrolyte membrane (PEM) fuel cell power system. It was found that the largest irreversibility rates took place in fuel cell stack. Both energy and exergy efficiency increased with the increment of stack operating temperature and operating pressure. However, the authors did not perform economic analysis on the findings.

Kamate and Gangavati (2009) presented exergy analysis of a heat-matched bagasse-based cogeneration plant in sugar factory. From exergy analysis, backpressure steam cogeneration plant has higher exergetic efficiency configuration than condensing steam turbine cogeneration plant. Boiler was the least efficient component while turbine was the most efficient component of the plant. The method to improve exergetic efficiency of boiler was to use higher high pressure steam inlet condition. Nevertheless, the authors did not discuss the economic potentials of using the higher pressure steam in the system.

2.2 Optimization of Scheduling Problems

The development of systematic techniques such as heuristic reduction methods, decomposition or aggregation techniques highlights the application of different optimization approaches. Many researchers have developed remarkable advances in modelling techniques, algorithmic solutions and computational technologies to solve scheduling problems.

Smaili et al. (1999) developed scheduling algorithm for the optimum cleaning of heat exchanger networks. The application of the technique was illustrated with a case study of the raw juice preheat train in a sugar refinery. Heavy fouling was found to occur in several of the heat exchangers using higher temperature vapour stream to achieve raw juice target temperature. The optimal scheduling was aimed at minimizing the consumption of higher temperature utility. This case study demonstrated that it was possible to use the fouling data gathered using plant performance systems to construct near-optimal cleaning schedules. The optimization problem formulation results in a NLP/MINLP solution pathway which proved to be robust and reliable. The non-convexity of the problem resulted in a number of sub-optimal but adequate solutions.

Georgiadis et al. (2000) presented optimal cleaning schedule of complex heat exchanger networks under fouling. The general fouling profiles was considered to predict the behaviour of the heat exchanger with time. It was shown that trade-offs existed between the total number of cleaning and timings of cleaning operations and the cost and availability of hot utility. The authors proved that the hot utility availability in the plant can have significant effect on the network cleaning policies. However, the model formulation cannot use for problems with more than 10 heat exchangers and the model is only reliable for short time horizon.

Pinto et al. (2000) discussed planning and scheduling applications for refinery operations. Firstly, the authors presented the development of a nonlinear planning model for refinery production. Real-world applications were developed for the planning of diesel production in the local refinery in Cubatao, Brazil. The authors reported that the new operating point managed to increase several million dollars in annual profitability. The second part of this work addressed scheduling problems in oil refineries that are formulated as mixed integer linear model. Their work considered the solution of optimization model for short term scheduling of a set of operations including product receiving from processing units, storage and inventory management in intermediate tanks, oil blending, and transport sequencing in oil pipelines.

In a subsequent work by Smaili et al. (2001) the mitigation of fouling in refinery heat exchanger networks was approached, using optimal management of cleaning. The results demonstrated expected features such as the regular cleaning of heat exchangers with significant impact on overall network performance and the order of cleaning actions.

Smaili et al. (2002) later developed optimization techniques for long term scheduling of heat exchanger network cleaning. This work explored the backtracking threshold accepting algorithm (BTA) approach. Two large scale case studies were considered involving 14 and 25 heat exchanger units. Both were tackled by outer approximation (OA) approach and BTA approach. From their result, BTA proved to be able to handle larger problems with more time intervals or exchangers than the commercial OA solver.

Alle et al. (2004) addressed the cyclic scheduling of cleaning and production operation in multiproduct multistage plants with performance decay. The authors reformulated their model by linearizing the nonconvex equations. The only nonlinear equation is the convex inequalities. The convex mixed integer non linear programming (MINLP) model was developed and solved to global optimality. The optimal cleaning sequence was produced with 4.9% reduction in cost compared to an alternate hierarchical solution.

Lavaja et al. (2004) proposed a new mixed integer linear model for the planning of heat exchanger cleaning in chemical plants. The authors developed decomposition procedure to decrease computational time. The model used two different fouling models which were linear fouling and asymptotic fouling model. The comparison was made between their solutions with moving horizon strategy, heuristic strategy and cyclic schedule. The authors found that moving horizon and heuristic strategy was not effective in solving this kind of problem. Meanwhile, cyclic schedule was not applicable to the model.

A year later, Lavaja et al. (2005) extended their previous model for the consideration of throughput reduction when the maximum capacity of the furnace was reached. The results showed that the only possibility for operating the network under critical conditions is by reducing the throughput during short periods of time when some of the heat exchangers must be put offline to be cleaned and the furnace cannot handle the required load. Reducing throughput in small percentages during short period of time allows for more cleaning and for recovering the performance of the network.

Another cleaning schedule for heat exchanger network was developed by Markowski et al. (2005). This work presented MINLP model formulation with the influence of fouling on heat exchangers. It was found that, the optimal cleaning schedule was able to save 5% of the maximum attainable value of energy recovered in heat exchanger network.

Castro et al. (2005) presented two alternative decomposition approaches for the efficient solution of large industrial scheduling problems. The first approach froze the schedule of the pre-assigned orders and ensured feasibility for the remaining through machine release dates. The second approach allowed for more flexibility by only fixing the assignments and relative positions of the previously scheduled orders. The authors found that the second approach was more robust and seems better suited the solution.

Sanaye et al. (2007) performed simulation of heat exchanger networks for estimating the amount of fouling, variations in overall heat transfer coefficient and variation in outlet hot and cold temperatures and planning the optimum cleaning schedules. The result showed that fouling decreased the overall heat transfer coefficient about 19% which results in about 4.6% decrease in heat transfer rate between hot and cold fluids. The optimal cleaning schedules for Urea and Ammonia units were found by minimizing operational cost using a numerical search method.

Ishiyama et al. (2007) presented the effect of fouling on heat transfer, pressure drop and throughput in refinery crude preheat train and the optimization of cleaning schedule. This work explored the interaction between thermal and hydraulic effects by using engineering analyses and fouling rate law. The authors compared thermal and hydraulic impacts on fouling of heat exchanger and found that thermal impact was more significant than hydraulic impact by comparing both impacts on different values of thermal conductivity of deposit. The authors also developed three case scenarios for crude preheat train cleaning schedule. From their result, manipulation of flow split during cleaning schedule represented reasonably cheaper than cleaning schedule without flow split. In addition, the improvement in performance was noticeable for coil inlet temperature (CIT) and throughput. Nevertheless, the model involved complex mathematical formulation.

In a recent work by Lavaja et al., (2009) a model for multiperiod cleaning under uncertainty with financial risk management was proposed. The model considers uncertainty in the future price of the natural gas expended in the furnace, the actual value of the fouling rates of the crude oil and the schedule of change of feedstock during the operation. The results show how the optimal strategies can vary when uncertainties are considered and how the model determines the best strategies to apply when risk is involved.

Xiau et al. (2010) presented a novel methodology for simultaneously optimizing synthesis and cleaning schedule of flexible heat exchanger network by genetic/simulated annealing algorithms (GA/SA). The authors introduced some effective strategies to reduce computational complexity. The first strategy was the use of T-H diagram approach to simplify the multi period synthesis of heat exchanger network. The second strategy was the simultaneous optimization can identify the most promising matches. The last strategy was the implementation of a novel continuous time representation of the scheduling problem. From their numerical example, it showed that the annual cost has reduced about 4.25%.

2.3 Reduction of Carbon Dioxide Emission in Industry

The concept of carbon credits came into existence as a result of increasing awareness on the need for pollution control. It was formalized in the Kyoto Protocol between 169 countries under the United Nations Framework Convention on Climate Change (UNFCCC). The Clean Development Mechanism (CDM) is the market-based mechanism of the Kyoto Protocol that involves non-Annex I countries to aim at promoting co-operative measures between Annex I or developed countries and the Non- Annex I or developing countries. The CDM is proposed with the twin objectives of helping Annex I countries to achieve their emission reduction targets and at the same time helping non-Annex I countries to promote sustainable development in their economies. The greenhouse gas reduction achieved by each CDM projects will be quantified in standard units known as Certified Emission Reductions (CERs) which is a form of carbon credit. It involves trading of emission reductions resulted from CDM project to countries that can use these CERs to meet their targets. There will be a transfer of money to the project that actually reduces the greenhouse emissions (Oh et al., 2010). The CDM project can be carried out in a number of sectors such as renewable energy, energy efficiency, transportation, agriculture or land use.

Shimazaki et al. (2000) developed a model analysis of “double clean development mechanism” to reduce both carbon dioxide and sulphur dioxide emissions in Japan and China. They found that double clean development mechanism could effectively mitigate the damage caused by sulphur dioxide emissions. Furthermore, it can also enhance the economies for both nations. However, China might not be able to maintain high growth rates under CO_2 and SO_2 restriction.

Gielen et al. (2002) analyzed the possible options for reduction of carbon dioxide emission in the Japanese petrochemical industry. The options were the introduction of biomass feedstocks, the introduction of new catalytic production processes and changes in wastes handling. From their result, about 4.9% emission reduction can be achieved by using biomass feedstock. Another 2.8% emission reduction can be achieved by changes in waste management. They also mentioned about uncertainties that may exist such as the availability of biomass and the feasibility of the technologies. This sector required further technology development to reduce more CO_2 emission.

Gadalla et al. (2006) presented estimation and reduction of carbon dioxide emissions from crude distillation units. The operating conditions of the existing crude oil distillation tower were changed using optimization method to reduce carbon dioxide emission to the minimum. In addition, integration of a gas turbine with the process furnace was proposed as a design option. The existing crude oil units can cut down their emissions by up to 22%. However, these units can achieve an emission reduction of up to 48% when a gas turbine is integrated. Both the operating costs and energy consumption were significantly reduced for relatively little capital investment and low payback time.

Gilau et al. (2007) addressed the cost effectiveness of renewable energy technologies in achieving low costs and promoting sustainable developments under the CDM. According to their results, hybrid systems of photovoltaic-diesel (PVDB), wind-diesel (WDB), and photovoltaic-wind-diesel (PVWDB) are very cost effective energy options compared with a diesel only energy option. The removal of about 87% carbon dioxide emissions could be achieved at negative cost but initial investment could increase by a factor of 20, which was one of the primary barriers hindering wider renewable energy applications in developing countries. Thus, the authors suggested CDM policy should investigate how to facilitate renewable energy projects through barrier removal in order to increase developing countries' participation in the carbon trading market. However, the proposed improvement options required high investment cost.

Purohit et al. (2008) analyzed CDM potential of solar water heating systems in India. From their estimation, the annual CER potential of solar water heating systems in India could theoretically reach 27 million tonnes. Annual CER capacity by 2012 could reach around 4-9 million tonnes and by 2020 it may be possible to reach 15-22 million tonnes. The limitations of their study were high capital cost and maintenance cost and also the solar technology was quite unfamiliar to the society.

Schneider et al. (2008) analyzed the CDM contribution to technology transfer. Technology transfer has been defined as a broad set of process covering the flows of know-how, experience and equipment. The authors presented three key findings concerning the CDM's role in contributing to technology transfer. First, the CDM increased the commercial viability of low-carbon technology transfer by setting a price on carbon. The high and stable carbon price would be desirable if technology transfer is to be increased under the CDM. Second, CDM reduced the barriers of lacking information and lacking access to capital on technology transfer. A database consists of information on the technological specification could be created by technology provider. The database can be implemented with CDM to serve valuable information for private sectors for choosing the proper technology and its provider. It also could serve the basis for better risk assessment of different technologies, thereby increasing access to capital. Third, the CDM does not improve the institutional framework of receiving countries which is considered vital to attracting international technology.

Show et al. (2008) presented emission reduction in anaerobic wastewater treatment. The project activity contributed to greenhouse emission reduction by recovering methane gas generated from anaerobic granulation process and utilising the biogas generated to displace fossil fuels for boilers. From their calculations, the proposed project activity required US$ 1.5 million with a reasonably attractive investment return period of less than 5 years by including revenue from the sale of CERs.

In a recent study, Kafle et al. (2009) reported feasibility study of capturing carbon credit benefits in an academic institution. The case study was the campus in India. The CDM opportunities identified in the hostels were solar water heating system, solar steam cooking system and energy efficient lighting and air conditioning system. From their results, solar steam cooking had negative value of internal rate of return (IRR) which was not commercially attractive. In the case of energy efficient lighting and air conditioners, these options did not require CDM revenue because IRR without CDM were already high. The only options required CDM benefit was solar water heating system because the IRR with CDM revenue was significant than IRR without CDM revenue. However, the drawback of their study was high initial cost and the consumers were less awareness about the benefit of capturing carbon credits.

Liu et al. (2009) discussed about long-term carbon storage and reduced carbon fluctuation that can be achieved by combination of improved forest management and efficient transfer of carbon into wood products. From their modelling result, they found that an important way to reduce global carbon emissions was through sustainable forest management.

Gaast et al. (2009) discussed an approach to facilitate low-carbon energy technology transfer to developing countries through the CDM. The objective of CDM is to support developing countries in achieving a sustainable path at the same time assisting developed countries in achieving their Kyoto Protocol. In the actual practice, CDM technology transfers are largely focused on reducing green house gas emissions at lowest cost and fewer needs on the domestic development as well as priorities of the developed countries. They figured out that new knowledge such as low-carbon technology transfer had insufficient information on what sustainability benefits can be delivered and how this can be achieved. Based on their study, they proposed to extend the technology experience by assisting more programs of low-carbon energy demonstration projects. This would improve awareness and trust in new low-carbon technologies.

Oh et al. (2010) conducted a review on the various energy efficiency efforts and carbon trading potential in Malaysia. A biomass project in Sabah, a state in east Malaysia, was the first carbon trading awarded CERs by the United Nation Executive Board of CDM. The corporate sectors such as power manufacturing, waste management, forestry, oil and gas manufacturing, agriculture and transportation sectors were identified as potentials beneficiaries to involve in CDM. As in March 2009, based on data released by United Nations Environment Programme (UNEP) resource centre, Malaysia had 156 projects or 4% of the total projects. As of March 2008, Malaysia already had 22 registered CDM projects with most the CERs coming from biomass plants. As of March 2007, two of the 22 CDM projects had sold 320,000 tonnes of CERs valued at less than RM 10 million.

2.4 Application of Alfa Laval Compabloc Heat Exchanger in Industrial Plant

De Castro et al. (2008) reported about a Compabloc unit that was positioned before a desalter unit in crude preheat train. The hot stream was vacuum residue. The tests were conducted for a range of throughputs over a period of 15 months. The thermal performance and fouling were investigated. It is noted that the maximum temperature of the heated crude is around 120^{o}C to 130^{o}C while the hot vacuum residue was around 200^{o}C. For shell and tubes heat exchanger, fouling could be expected on both sides of heat exchanger. For heat exchanger that was positioned before the desalter, it was common to find deposits with high concentration of inorganic salt, iron and sulphur. However, no fouling was observed for Compabloc heat exchanger under any of these test conditions. From the test result, the user decided to install more Compabloc in the preheat train. They also had an interest to carry out further performance and fouling tests at higher crude temperature and with partial crude vaporization.

Spangler et al. (2006) reported on a Compabloc exchanger installed at the hot end of crude preheat train. This unit extracted heat from heavy vacuum gas oil (HVGO) pump around stream. After five months of operation, the unit was shutdown for cleaning because the exchanger was partially blocked with particulate material due to the problem in the existing piping that had not been flushed. Once the exchanger had been cleaned, satisfactory operation was obtained throughout the normal operating period. In fact even after eighteen months continued operation, the performance of the unit exceeded the original design objective.

Arvidsson (2003) mentioned about the application of Compabloc heat exchanger in Yukos oil refinery in Syzran, Russia. When a new part of the plant was built, the Compabloc was installed instead of traditional shell and tube units, both as a reboiler and as a condenser. The Compabloc heat exchangers were installed at a stripper column in a gas sweetening process. The decision to install Compabloc instead of traditional shell and tube heat exchanger enables the plant to save both money and space. The Compabloc heat exchangers had been in operation since January 2002.

Alfa Laval (2002) published a case story about the installation of Compabloc condenser instead of shell and tube exchanger as the overhead condenser of fluidized catalytic cracking (FCCU) fractionator at the Shell refinery in Sarnia, Ontario Canada. The objective of installing new Compabloc was to recover significant amount of heat in the FCCU into the treated cold water flow used as make-up water for the steam plant. Furthermore, the Compabloc condenser was made by Hastelloy C276 that completely eliminated the corrosion problem. The corrosion was previously occurred in tube bundle of shell and tube exchanger caused by bisulphides, chlorides and cyanides that were presented in overhead vapour.

Another case story published by Alfa Laval corporate website is the application of Compabloc at Queensland Alumina Plant in Australia. In 2001, the engineers decided to replace a shell and tube unit with Alfa Laval Compabloc unit in order to decrease maintenance cost. The excessive level of fouling in the original shell and tube exchanger need regular cleaning every 6 months. The cleaning procedure took a week to complete. The shell and tube exchanger also suffered from internal leakage due to extensive corrosion. The installation of Compabloc reduced fouling problems and corrosion which resulted to significant service cost reduction. After 12 months of the installation, the Compabloc is in successful operation without having to be cleaned once.

2.5 Summary of Literature Review

The application of exergy analysis in industry has clearly indicated the locations of exergy inefficiencies in the processes and highlighted the potential improvements of the processes. The reviews on the application of exergy analysis in industrial plant with its improvement options and limitations have been discussed. From exergy analysis, they managed to identify process inefficiencies in the systems and proposed the options to improve the systems. However, not many of them have conducted economic analysis on the proposed improvement options to show the reliability of the improvement options on capital and operational cost.

Meanwhile, literatures in scheduling area have highlighted the successful application of different optimization approaches to an extensive variety of challenging problem. The reviews on optimization of cleaning schedule in heat exchanger network with implementation of various optimization methods, the advantages and limitation of their cleaning schedule model has been discussed. The common problem formulation is Mixed Integer Non Linear Programming (MINLP) and Mixed Integer Linear Programming (MILP) that can be solved by using readily available optimization software.

Many studies mainly in energy sectors have been done by previous researchers to generate energy improvement options in order to reduce carbon dioxide emission. The implementations of energy improvement options to minimize carbon dioxide emissions are crucial due to some barriers such as high capital cost, lack of commercial viability and lack of awareness from the communities. However, these barriers may be overcome by introducing the concept of carbon credit in the economic analysis of energy improvement options under CDM project. The reviews about carbon dioxide emission reduction project from different sectors with respective emission reduction and a few limitations have been briefly discussed. The energy efficient technologies will be more feasible, attractive and cost effective with implementation of carbon credit.

The Alfa Laval Compabloc heat exchanger has been installed in numerous refineries around the world in various process units. Compabloc is a welded plate heat exchanger without gasket that can be used in high pressure and high temperature duties in aggressive media. Compabloc are highly compact heat exchanger that can overcome many space restrictions. The high turbulent flow created in corrugated pattern in Compabloc minimizes fouling problem due to high film coefficient and high shear wall stress. The reviews about the application of Compabloc in industrial plant have been explored.

CHAPTER 3

METHODOLOGY

3.1 Introduction

Exergy Analysis is conducted to establish the benchmark of location and quantity of exergy losses as well as to identify improvement options of the plant. The agreed process data is simulated in Process Integration ‘WORK’ software to obtain exergy composite curve. Exergy composite curve is a graph of hot and cold composite curves on a carnot factor (Y) versus enthalpy (X) axis.

The fuel reduction strategies are generated from exergy composite curve analysis. The proposed options are reduction of heat loss from furnace stack by reducing stack temperature and enhancement of heat recovery in the CPT by improving furnace inlet temperature.

For second option, there are three cases namely plant shutdown for crude preheat train overall cleaning (Case 1), heat exchangers online cleaning (Case 2) and heat exchangers online cleaning with retrofit of high efficiency heat exchangers (Case 3). All the models are coded in the commercial optimization software, GAMS version 23.2. Mixed Integer Non Linear Programming (MINLP) cleaning schedule optimizer models are applied for all the cases.

The carbon credit or carbon emission reduction (CER) revenue is calculated for all the options in economic analysis. CER are carbon credit issued by the Clean Development Mechanism (CDM) Executive Board achieved by CDM projects and verified by a Designated Operation Entity (DOE) under the rules of the Kyoto Protocol.

3.2 Establishment of Base Case Data

The relevant process data is obtained from operational, simulation and design data. The selected data has been agreed and validated by a local oil refinery's engineers after careful consideration. Temperatures of hot and cold streams of CPT are obtained from manual temperature monitoring spreadsheet prepared by refinery production department. The process flow rate was taken from process information (PI) historical data over 12 months while the material properties for crude oil and its products are obtained from the simulation data prepared by the plant engineers. The data extracted from CDU is attached in Appendix A.

Figure 3.1 shows a simplified process flow scheme of CDU crude preheat train. From the process flow scheme, the hot streams are pre-flash vapour, kerosene, diesel, LSWR, kerosene pump around, diesel pump around and pre-flash bottom. The hot stream for heat exchanger E2 and E9 is kerosene exit. The hot stream for heat exchanger E3, E7 and E10 is diesel exit. The hot stream for heat exchanger E4, E11 and E12 is LSWR exit.

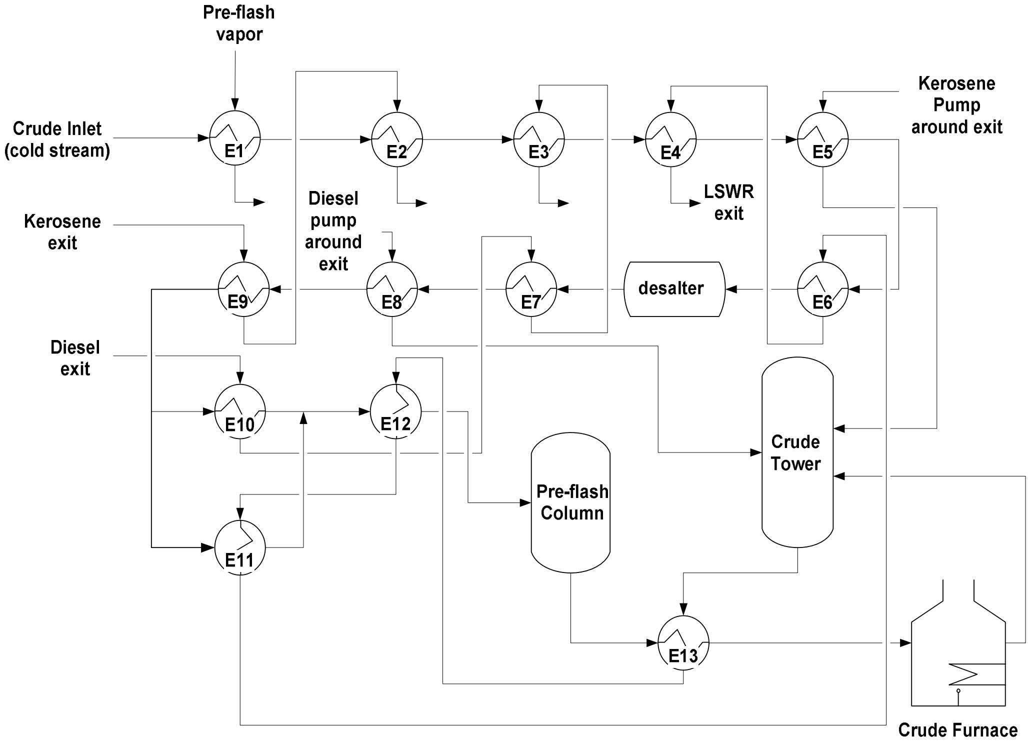

Figure 3.1 CDU crude preheat train flow scheme

The crude oil is heated up to 112^{o}C before entering the desalter. Table 1.1 lists heat exchangers before desalter unit with hot streams for heat recovery. The hot streams are the hot product streams coming out from the CDU. After desalter, the crude oil stream is further heated up using heat recovered from the process streams. The list of heat exchangers with hot streams after desalter unit is shown in Table 1.2. At this end, the crude oil starts to vaporize at 203^{o}C. Then, the crude oil enters a pre-flash column to remove light naphtha, mixed naphtha and light hydrocarbon gases from the crude oil. The pre-flash column is equipped with 18 single passes fractionation trays located above the crude inlet. The vapour is risen up to pre-flash overhead distillate and the liquid flows downward to the bottom. The preflash column bottom is further heated by heat exchanger E13 before entering furnace at design temperature of 215^{o}C.

Table 3.1 Heat exchangers before desalter unit

Heat exchangers	Hot streams
E1	Pre-flash overhead exit
E2	Kerosene exit
E3	Diesel exit
E4	LSWR exit
E5	Kerosene Pump Around exit
E6	LSWR exit

Table 3.2 Heat exchangers after desalter unit

Heat exchangers	Hot streams
E7	Diesel exit
E8	Diesel pump around exit
E9	Kerosene exit
E10	Diesel exit
E11	LSWR exit
E12	LSWR exit
E13	Pre-flash bottom exit

Additional process heating is provided by crude furnace. The crude oil enters four passes of furnace, which is heated by the fuel gas burners. The heated crude oil exits the furnace at approximately 367°C before entering the CDU column flash zone. The CDU column is equipped with 29 valves type trays above the flash zone and 4 valve type trays below the flash zone. The vapour from the flash zone at the top section ascends the trays where they are fractionated into diesel, kerosene, heavy naphtha and overhead distillate. Meanwhile, the bottom section of the column is low sulphur waxy residual (LSWR). There are three refluxes which consist of overhead reflux naphtha, kerosene pump-around and diesel pump around.

3.3 Exergy Analysis

Figure 3.2 shows an exergy composite curve. The upper line is the hot composite curve and the lower line is the cold composite curve. The area under the hot composite curve is the amount of exergy source (ΔE_H) and the area under the cold composite curve is the amount of exergy sink (ΔE_C). Note that ΔE_H is partly covered by ΔE_c in Figure 1. The gap between hot composite curve and cold composite curve is the exergy loss which is $\Delta Ex_{loss} = \Delta E_H - \Delta E_C$.

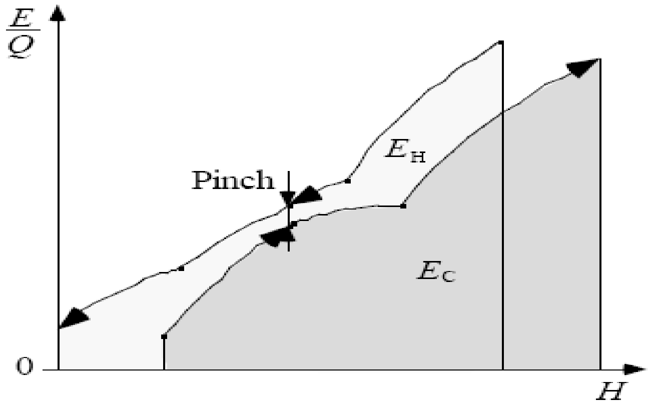

Figure 3.2 Exergy Composite Curve

The exergy source in a system is provided by hot process streams that transfer heat and is calculated as follows:

$$\Delta Ex_{source} = (H_{h2} - H_{h1}) - T_o(S_{h2} - S_{h1}) \tag{3.1}$$

On the other hand, the cold process stream that receives heat is the exergy sink:

$$\Delta Ex_{\sin k} = (H_{c2} - H_{c1}) - T_o(S_{c2} - S_{c1}) \tag{3.2}$$

Exergy is never conserved in real processes. Exergy will degrade and will be lost. Exergy loss reflects the irreversibility in the heat transfer process. Exergy loss can be calculated from an exergy balance as follows:

$$\Delta Ex_{loss} = \sum \Delta Ex_{sources} - \sum \Delta Ex_{\sin ks} = T_o(\Delta S_c - \Delta S_h) \tag{3.3}$$

where ΔS_c is change in entropy for cold streams and ΔS_h is change in entropy for hot streams.

Since there are difficulties to obtain thermodynamic data for every process especially entropy, an equation has been developed which only need the information about pressure, temperature and enthalpy of the process streams. For every liquid and gas at constant pressure, the enthalpy and entropy change are:

$$\Delta H = H_2 - H_1 = m\int C_p dT \tag{3.4}$$

$$\Delta S = S_2 - S_1 = m\int C_p \frac{dT}{T} \qquad 3.5)$$

Assume that the specific heat, Cp is constant for an ideal gas and incompressible liquid. Equations (3.4) and (3.5) are integrated to gives

$$\Delta S = \Delta H \left| \frac{\ln\left(\frac{T_2}{T_1}\right)}{T_2 - T_1} \right| \tag{3.6}$$

Substituting equation (3.6) into equation (3.3) gives exergy loss equation (3.7)

$$\Delta Ex_{loss} = T_o \Delta H \left(\frac{\ln\left(\frac{T_{c2}}{T_{c1}}\right)}{T_{c2} - T_{c1}} - \frac{\ln\left(\frac{T_{h2}}{T_{h1}}\right)}{T_{h2} - T_{h1}} \right) \tag{3.7}$$

From exergy loss equation in equation (3.7), the amount of exergy loss for process equipment is calculated to determine which equipment has the highest exergy loss.

The extracted process data from the refinery are simulated in WORK software to obtain exergy composite curve. The enthalpy of process streams are calculated by using the following equation.

$$\Delta H = mC_p \Delta T \tag{3.8}$$

where m is mass flow rate, C_p is specific heat and ΔT is the temperature difference. Table 3.3 shows the properties to calculate enthalpy of process streams.

Table 3.3 Properties to calculate enthalpy of process streams

Properties	Values
Specific heat for crude oil	2.0042 kJ/kg^{o}C
Crude oil flow rate	75 kg/s
Preflash bottom flow rate	68 kg/s

The analysis of exergy loss in furnace requires the determination of theoretical flame temperature (TFT). TFT is the highest achievable temperature provided by combustion of burning fuel. The typical fuel for combustion in the furnace is the off gas as listed in Table 3.4. The main assumption is that no heat is lost from the combustion system. It is assumed that nitrogen does not involve in combustion system since nitrogen is an inert gas.

By assuming the combustion system is adiabatic with excess air of 10%, TFT is calculated using energy balance as in equation (3.9). The calculation of TFT is shown in Appendix B. An example about the calculation of TFT is provided below.

$$\int_{T_{inlet}}^{T_o} \Delta H_{reactant} + \Delta H_C^0 + \int_{T_0}^{TFT} \Delta H_{product} = 0 \tag{3.9}$$

Table 3.4 Typical fuel for combustion

Off gas	Amount (mol%)
hydrogen	52.60
methane	29.76
ethane	8.75
propane	5.63
n-butane	1.01
hexane	0.98
i-butane	0.94
i-pentane	0.16
n-pentane	0.16

Example: TFT calculation for methane fuel.

Methane is burnt with 10% excess air. Methane and air is fed at 25°C. The standard heat of combustion ($\Delta H^o c$) for methane is -802,600 kJ/ kmol. Nitrogen, N_2 does not involve in reaction. The combustion reaction is:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \tag{3.10}$$

The reactants are CH_4, O_2 and N_2. The amount of reactants is calculated by referring to equation (3.10). From the equation, 1 kmol of CH_4 is reacted with 2 kmol of O_2 to produce 1 kmol of CO_2 and 2 kmol of H_2O. Thus, the amounts of reactants in kmol are,

CH_4 = 1

O_2 = 2 (1.10) =2.2

N_2 = 2.2(79/21) = 8.65

For O_2, 10% excess air is considered by multiplying with 2 kmol. For N_2, ratio of N_2/O_2 in air composition is multiplied with 2.2 kmol of O_2.

The amounts of products in kmol are,

CO_2 = 1

H_2O = 2

Excess O_2 = 2.2 – 2.0

= 0.2

Inert N_2 = 8.65

TFT is assumed at 1800 ^{o}C. The specific heat, Cp for CO_2, H_2O, N_2 and O_2 are 54.18 kJ/kmol K, 42.87 kJ/kmol K, 34.96 kJ/kmol K and 33.18 kJ/kmol K. TFT is calculated by using energy balance as in equation (3.8). The combustion system is adiabatic where T_o is equal to T_{inlet} for $\Delta H_{reactant}$. Thus, $\Delta H_{reactant}$ is zero. The equation becomes:

$$\Delta H_C^0 + \int_{T_0}^{TFT} \Delta H_{product} = 0 \qquad (3.11)$$

The integration of $\Delta H_{product}$ gives,

$\Delta\ H_{product}$ = [54.18 + 2(42.87) + 0.2(34.96) + 8.65(33.18)] [TFT-298]

= 433.92 (TFT-298)

433.42 (TFT-298) = 802,600

TFT = 2150 K ≈ 1877^{o}C

The calculated value of TFT is compared with the assumed value of TFT. The percentage of error difference should be less than 5%. The percentage of error for this example is 4%. Thus, the value of calculated TFT is acceptable.

3.4 Generation of Possible Fuel Reduction Strategies

Figure 3.3 shows a typical exergy composite curve with the location of TFT, T stack, TIT and FIT. T stack is located at the initial point of hot composite curve while TFT is located at the end point of hot composite curve. FIT is located at the initial point of cold composite curve while TIT is located at the end point of cold composite curve.

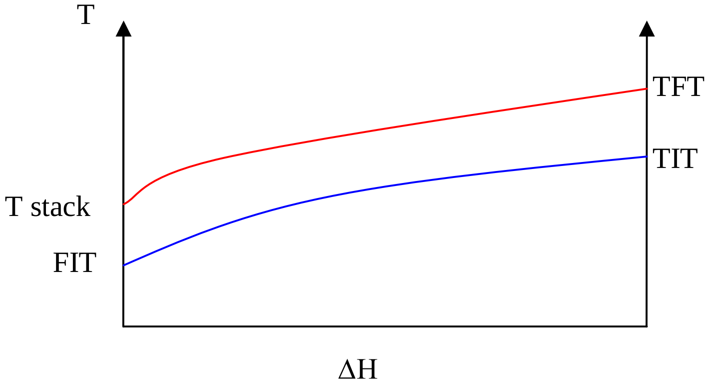

Figure 3.3 Location of TFT, T stack, FIT and TIT on exergy composite curve

The fuel reduction strategies are generated from exergy composite curve analysis. As shown in Figure 3.3, amount of exergy loss is represented by the gap between hot and cold composite curve. Thus, the idea to minimize exergy lossch of the system is able to obtain closer gap between hot and cold composite curve. The closer gap between hot and cold composite curve can be obtained by:

a) Reducing Theoretical Flame Temperature (TFT)

b) Reducing Stack Temperature (T stack)

c) Increasing Tower Inlet Temperature (TIT)

d) Increasing Furnace Inlet Temperature (FIT)

These four choices are the possible fuel reduction strategies for CDU. From these choices, two choices are chosen to be implemented in this study which are reducing T stack and increasing FIT. Thus, the proposed options for fuel reduction strategies in this study are reduction of heat loss from furnace stack by reducing T stack and heat recovery from CPT by improving FIT.

3.5 Model Formulation

The proposed fuel reduction strategies to be implemented in CDU are reduction of heat loss from furnace stack by increasing heat recovery of convection section of the furnace and heat recovery from CPT by improving FIT. Development of the model formulation for reduction of heat loss from furnace stack and heat recovery from CPT as well as the calculation for carbon dioxide emission are explained below.

3.5.1 Reduction of Heat Loss from Furnace Stack

The process fluid goes to convection section and then flows to radiation section of the furnace. The operation will control the opening of stack damper in order to recover more heat at the convection section. However, during this adjustment, the amount of excess oxygen will be monitored closely to avoid incomplete combustion.

When more heat is recovered in the convection section, process fluid will have higher amount of heat that are going to radiation section. Thus, the firing of fuel gas consumption in the furnace will reduce due to lower stack temperature of furnace in convection section. Reducing stack temperature will recover the available heat instead of wasting the heat into the atmosphere.

The reduction of heat loss from furnace stack is demonstrated in Figure 3.4. The temperature versus enthalpy in furnace operation is shown. When T stack_1 is reduced to T stack_2, the amount of heat loss is reduced. From equation (3.12), η is the furnace efficiency, ΔT is the temperature difference from theoretical flame temperature (TFT) to stack temperature (T stack), m is flue gas flow rate, C_p is the specific heat of the flue gas and ΔH_c is the heat of combustion of the fuel gas.

$$\eta = \frac{Q_{process}}{Q_{fuel}} = \frac{mC_p\Delta T}{\Delta H_c} \tag{3.12}$$

From this equation, when T stack is reduced, ΔT will be increased. Thus, the furnace efficiency, η increases.

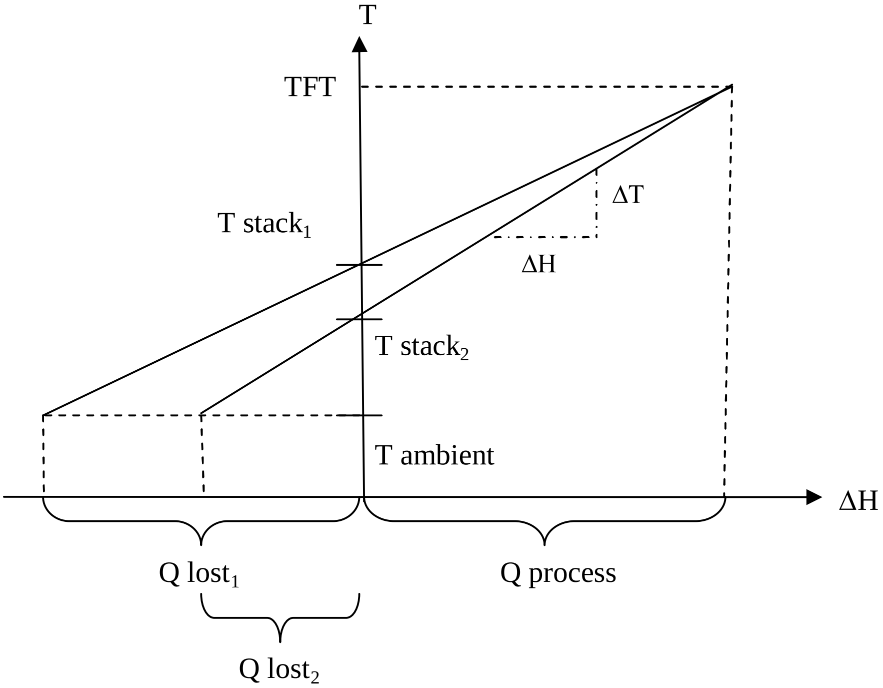

Figure 3.4 Temperature-enthalpy profile for furnace operation

Fuel gas flow rate is calculated by dividing furnace duty with heat of combustion of the fuel gas. This is the fuel gas consumption at 100% efficiency. The fuel gas consumption at 100% efficiency is divided by actual furnace efficiency to obtain the exact fuel gas consumption. Equation (3.13) shows the equation to calculate fuel gas flow rate.

$$m_{fuel} = \frac{Q_{furnace}}{\Delta H_c} \times \frac{1}{\eta} \tag{3.13}$$

3.5.2 Enhancement of Heat Recovery in the Crude Preheat Train

The method to enhance heat recovery in the CPT is by improving furnace inlet temperature (FIT). Three cases are proposed to improve FIT, namely plant shutdown for crude preheat train overall cleaning (Case 1), heat exchangers online cleaning (Case 2) and heat exchangers online cleaning with retrofit of high efficiency heat exchangers (Case 3).

The cleaning schedule formulation for Case 2 and Case 3 is incorporated by the following simplifications:

(i) Constant mass flow rates for all hot and cold streams
(ii) Constant linear fouling rates obtained from historical data
(iii) Pressure drop considerations were not included
(iv) Only heat exchangers in parallel position were allowed to perform online cleaning

All the models are coded in the commercial optimization software, GAMS version 23.2. Mixed Integer Non Linear Programming (MINLP) cleaning schedule optimizer models are applied for all the cases. The historical data for parameters such as fuel gas flow rate, furnace inlet temperature, hot streams and cold streams of heat exchanger network are collected. The historical data is needed to obtain correlation and profiles of the respective parameters.

3.5.2.1 Coding the Model in GAMS Software

General Algebraic Modelling System (GAMS) is a software product of the GAMS Development Corporation to solve mathematical problems of linear, integer linear, nonlinear and integer nonlinear programs.

The input components of GAMS software are known as sets, data, variables, equations, model and solve statements. The first input components are sets. Set is the indices in the algebraic representations of the model. For example, the set for Case 1 is months (*t*) as represented below.

```
SET
t months /0*59/;
```

In GAMS model, three different formats are allowable for entering data. The formats are lists, tables and direct assignments.

The third input component is variables. The variables of a GAMS model must be declared with variable statement. Each variable is given a name, a domain if appropriate and text. The Case 1 model contains the following variable statements as below.

```
POSITIVE VARIABLE

FIT(t)   Furnace inlet temperature
FG(t)   fuel gas consumption
CO2(t) carbon dioxide emission;

VARIABLE
z objective function minimize cost;

BINARY VARIABLE
 y(t) heat exchanger is cleaned at period t;
```

The forth input component in GAMS is equations statement. The format of the equation statement is the keyword equation followed by the name, domain and the text of equations or inequalities being declared. The Case 1 model contains the following equation declaration.

```
EQUATIONS
cost                 minimize cost
```

```
FuelFIT      FG vs FIT correlation
initialFIT   initial fuel gas
Fuel         fuel gas correlation
lowerFIT     lower FIT
CO2emission  carbon dioxide emission;
```

The components of an equation definition are the name of equation, domain, symbol '..', left hand side expression, operator and right hand side expression. The equation definition for Case 1 model contains six statements. The statements are:

```
Fuel(t)..            FIT(t)    =e= (FIT(t-1)-0.5603)*(1-y(t))+(FIT('0')*y(t));
FuelFIT(t)..         FG(t)     =e=  -2.2564*(FIT(t)*FIT(t))+902.58*FIT(t)-86575;
initialFIT('0')..    FIT('0')  =e= 215;
lowerFIT(t)..        FIT(t)    =g= 200;
CO2emission(t)..     CO2(t)    =e=1.32*FG(t);
cost..                         z =e= sum(t,(FG(t)-FG('0')))*0.04072*14.549*720 +
                     sum(t,y(t))*(520000+5450000);
```

The format to write model statement is the keyword 'model' followed by the name of the model and lists of equations names encloses in slashes. The lists of equations names can represent by shortcut /all/. For Case 1, the model statement is:

```
Model scheduling /all/;
```

The solve statement is written after the model statement. The format to write solve statement is the keyword 'solve' followed by the name of the model, the keyword 'using', the available solution programming, the keyword 'minimizing' or 'maximizing' and the name of objective function. The solve statement for Case 1 is:

```
Solve scheduling using MINLP minimize z;
```

The coding of the model for all the cases in Option 2 is attached in Appendix D.

3.5.3 Calculation of Carbon Dioxide Emission

The carbon dioxide emission is calculated by the following equation.

$$CO_2\,(\frac{kg}{h}) = Specific\ CO_2\ (\frac{kg}{kWh}) \times Heating\ value(\frac{kWh}{kg}) \times density\ fuel(\frac{kg}{m^3}) \times Fuel\ gas\ (\frac{m^3}{h}) \qquad (3.14)$$

Table 3.5 shows properties for each component in fuel gas. Density of fuel gas is calculated using equation (3.15) where P is pressure, M is molar mass, R is ideal gas constant and T is temperature. The value of P is 101.3 kPa, R is 8.3143 kPa.m^3/kmol.K, T is 300 Kelvin and M is 13.11 kg/kmol.

$$\rho = \frac{PM}{RT} \tag{3.15}$$

Table 3.5 Properties for each component in fuel gas

Component	Molecular weight	Composition (kmol%)	Molar mass for each component (kg/kmol)	Heat of combustion (kJ/kmol)
hydrogen	2.00	0.526	1.05	241,800
ethane	30.00	0.088	2.62	1,428,600
propane	44.00	0.056	2.46	2,043,100
i-butane	58.12	0.009	0.54	2,657,300
i-pentane	72.15	0.002	0.11	3,244,900
methane	16.00	0.298	4.76	802,600
n-butane	58.12	0.010	0.59	2,657,300
n-pentane	72.15	0.002	0.12	3,244,900
hexane	86.18	0.010	0.84	3,855,100

Table 3.6 shows the properties of natural gas and refinery fuel gas to calculate carbon dioxide emission. The heating value and density of refinery fuel gas are calculated. The value of specific CO_2 emission for refinery fuel gas is estimated. The estimation of specific CO_2 emission for refinery fuel gas is referred to specific CO_2 emission for natural gas. The density of refinery fuel gas is divided by density of natural gas and multiplies with specific CO_2 emission for natural gas to obtain specific CO_2 emission for refinery fuel gas.

Table 3.6 List of properties for natural gas and refinery fuel gas to calculate carbon dioxide emission

Properties	Natural gas	Refinery fuel gas
Specific CO_2 emission	0.200 kg/kWh (Volker, 2005)	0.166 kg/kWh
Heating value	17.644 kWh/kg (Perry et al., 1999)	14.95 kWh/kg
density	0.641 kg/m^3 (Perry et al., 1999)	0.530 kg/m^3

3.6 Feasibility and Economic Analysis

The feasibility and economic analysis are conducted to investigate the performance of fuel reduction strategies. Feasibility analysis is conducted to investigate whether all the options are viable and possible to implement in the plant. The criterion to decide the feasibility of options are comparison of energy and carbon dioxide saving and comparison of FIT and extra fuel gas (ΔFG) profiles for all the options with current practice performance. For second option, heat exchangers cleaning schedules for all the cases are presented in feasibility analysis section.

Meanwhile, the economic analysis is conducted to determine the operational cost for all the options. The operational cost for all the options are compared with operational cost for current practice to obtain percentage of cost saving. The operational cost is calculated with and without carbon credit revenue for all the options.

For Case 3, the purchase cost for high efficiency heat exchangers are calculated to obtain payback period. Seider et al. (2003) lists the purchase cost equation for chemical processing equipments. The purchase cost for plate and frame heat exchanger are as in equation (3.16) below. C_{ph} is the purchase cost (\$), A is the heat transfer area (ft^2) with the range of 150-15000 ft^2. The material is stainless steel. The operating pressures are limited to 300 psig.

$$C_{ph} = 7000A^{0.42} \tag{3.16}$$

Compabloc welded plate heat exchanger is the most highly efficient compact heat exchanger with design pressure up to 450 psig (Gunnarsson et al., 2006).

The proposed purchase cost equation is as follow.

$$C_{ph} = 7000A^{0.492} \tag{3.17}$$

This equation (3.17) is proposed by referring to the manufacturer’s budgetary proposal for purchasing Compabloc welded plate heat exchanger that was presented to a local refinery. From the proposal, the heat transfer area is 900.917 ft^2 and the purchase cost is $198,650. As in the above equation, the corrected coefficient for purchase cost equation is 0.492.

3.7 Summary of Research Methodology

The summary of overall research methodology is illustrated as the flow chart in Figure 3.5. As previously discussed in the above sections of this chapter, the first step in research methodology is the establishment of base case data. The base case data is verified and agreed by plant engineers. The base case data is used to perform exergy analysis. Exergy analysis is conducted to determine location and magnitude of exergy loss in the current practice system. The next step is to generate possible fuel reduction strategies from exergy composite curve analysis. The change in exergy loss per degree Celsius is calculated to determine whether the possible strategies are able to reduce fuel consumption in the furnace. There are two proposed options for fuel reduction strategies in this study which are reduction of heat loss from furnace stack and enhancement of heat recovery in the CPT. Model formulations for these options are developed. The models are solved by using optimization software to obtain optimal results. The results of the options are compared with the current practice by conducting feasibility and economic analysis to determine whether the options are feasible and cost-effective options.

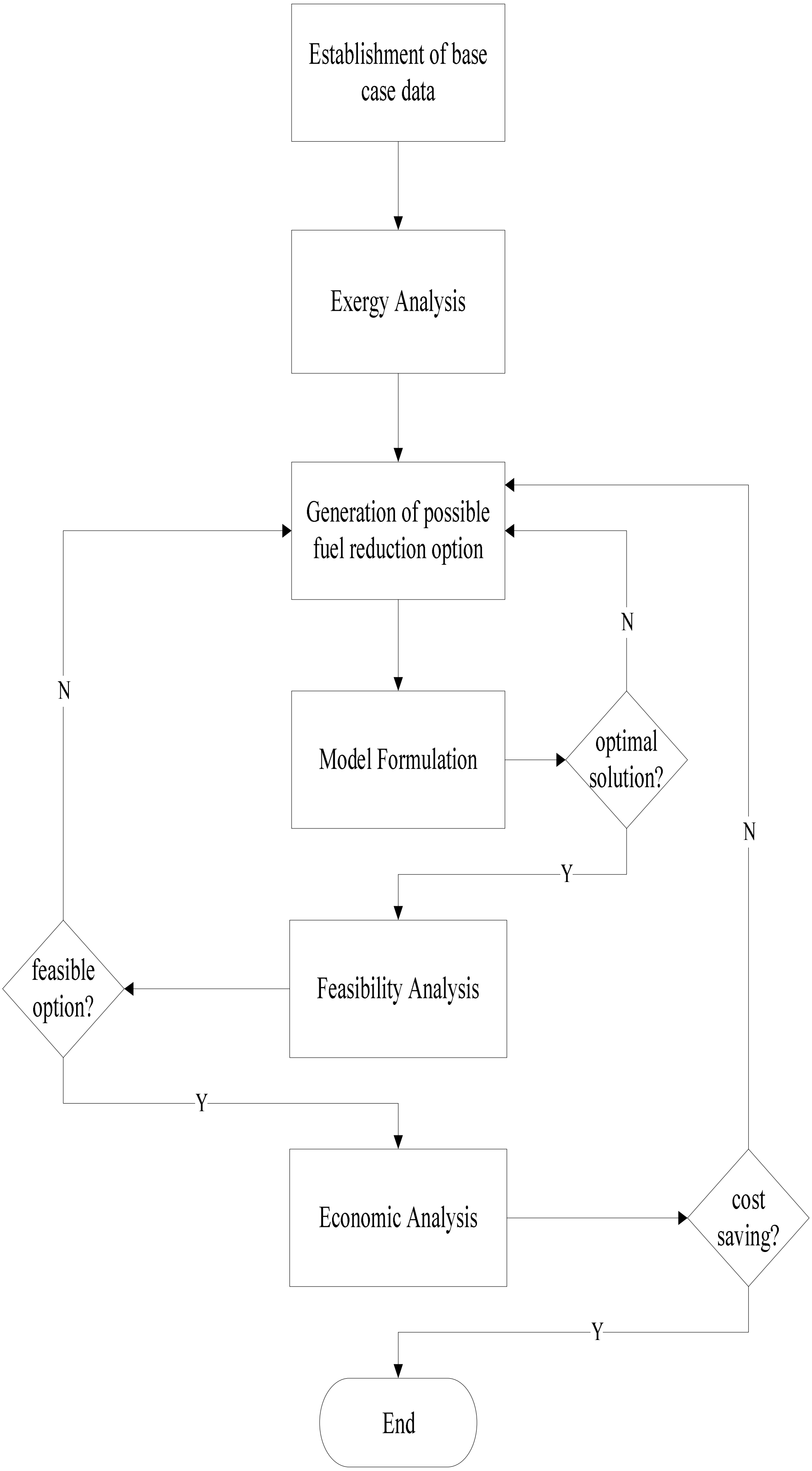

Figure 3.5 Flowchart of the overall research methodology

CHAPTER 4

RESULT AND DISCUSSION

4.1 Exergy Analysis

4.1.1 Exergy Composite Curve

The hot and cold streams of the process are described in the Table 4.1 and Table 4.2 respectively. From Table 4.1, stream C14 is the cold stream of furnace from furnace inlet temperature (FIT) to tower inlet temperature (TIT). Meanwhile, from Table 4.2, stream H14 is the hot stream of furnace from theoretical flame temperature (TFT) to stack temperature (T stack).

Table 4.1 Temperature inlet and outlet for cold streams

Streams	Cold streams	Ti (^oC)	To (^oC)	ΔH (kW)
C1	Crude oil to E1	33	66	4959.43
C2	Crude oil to E2	66	89	3456.57
C3	Crude oil to E3	89	100	1653.14
C4	Crude oil to E4	100	102	300.57
C5	Crude oil to E5	102	106	601.14
C6	Crude oil to E6	106	112	901.72
C7	Crude oil from desalter to E7	120	160	6011.43
C8	Crude oil to E8	160	200	6011.43
C9	Crude oil to E9	200	205	751.43
C10	Crude oil to E10	205	230	2098.50
C11	Crude oil to E11	205	215	663.46
C12	Crude oil to E12	223	232	1352.57
C13	preflash bottom to E13	200	215	2041.78
C14	FIT to TIT	215	367	50795.00

Table 4.2 Temperature inlet and outlet for hot streams

Streams	Hot streams	Ti (^{o}C)	To (^{o}C)	ΔH (kW)
H1	Preflash vapour to E1	83	40	4959.43
H2	Kerosene exit to E2	205	79	3456.57
H3	Diesel exit to E3	170	120	1653.14
H4	LSWR to E4	185	179	300.57
H5	kerosene pump around to E5	145	130	601.14
H6	LSWR to E6	225	185	901.72
H7	Diesel exit to E7	235	170	6011.43
H8	Diesel pump around to E8	205	180	6011.43
H9	kerosene exit to E9	230	205	751.43
H10	Diesel exit to E10	279	235	2098.50
H11	LSWR to E11	242	225	663.46
H12	LSWR to E12	300	242	1352.57
H13	LSWR from crude tower to E13	347	300	2041.78
H14	TFT to Tstack	1615	290	50795.00

Exergy composite curve is obtained from the simulation of extracted process data from the refinery by using WORK software. The ambient temperature (T_o) of the system is 27^{o}C.

The exergy composite curve for hot and cold composite is shown in Figure 4.1. The upper line is the hot composite curve and the lower line is the cold composite curve. The area under the hot composite curve is the amount of exergy source (ΔEx_{source}) and the area under the cold composite curve is the amount of exergy sink (ΔEx_{sink}). Exergy loss (ΔEx_{loss}) is the area between exergy source and exergy sink.

The axis of exergy composite curve is carnot factor (Y) versus enthalpy (X). Carnot factor (CF) is the measure of exergy level, CF = $1-T_o/T_i$, where T_o is the ambient temperature and T_i is the process temperature. The temperature unit to calculate carnot factor is Kelvin.

The region that contributes to the highest exergy is identified as the largest gap between hot and cold composite curves. In Figure 4.1, the largest gap is at the end of the curve which is at the temperature region of furnace. Furnace has temperature region from TFT to stack temperature for hot composite curve and temperature region from FIT to TIT for cold composite curve. The value of TFT to stack temperature is 1615^{o}C to 290^{o}C and the value of FIT to TIT is 215^{o}C to 367^{o}C. Table 4.3 demonstrates the value of carnot factor for respective temperature region of furnace. The approximate value of enthalpy for this region is 51000 kW.

Table 4.3 Carnot factor for temperature region of furnace

Temperature region	Values (K)	Carnot factor
TFT	1888	0.84
T stack	563	0.47
FIT	488	0.39
TIT	640	0.53

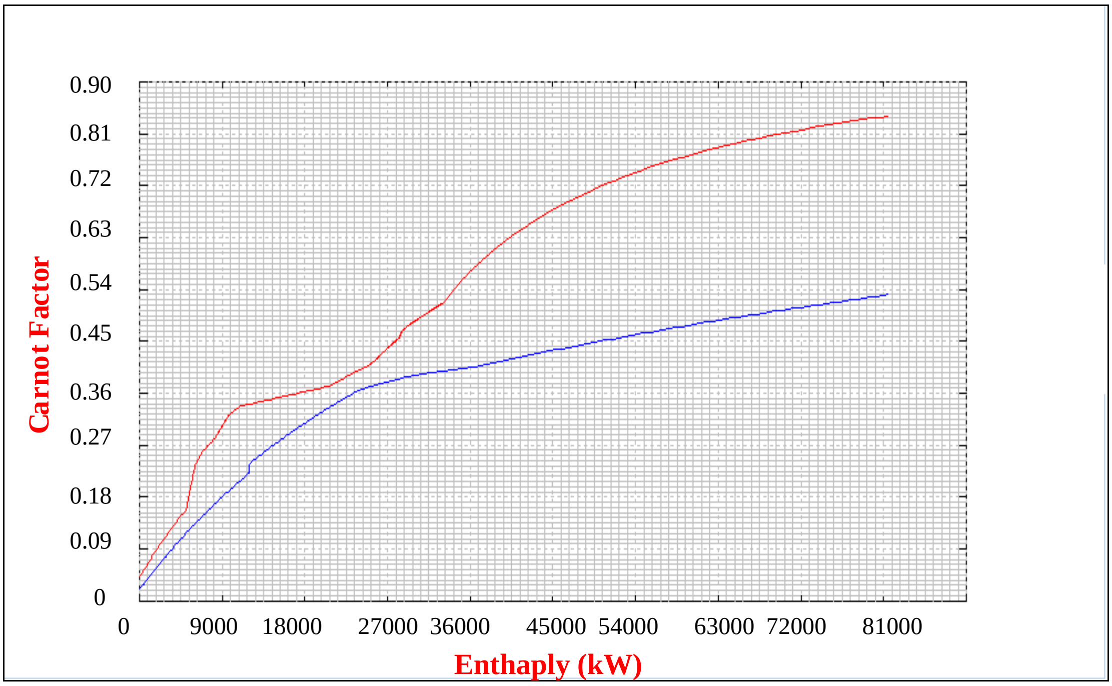

Figure 4.1 Exergy Composite Curve for CDU crude preheat train

4.1.2 Distribution of Exergy Loss

Table 4.4 shows amount of exergy source, exergy sink and exergy loss for heat exchanger network and furnace. From Table 4.4, furnace contributes some 86% of total exergy loss in the system.

Table 4.4 Exergy loss for heat exchangers and furnace in CDU crude preheat train

Furnace/Heat exchangers	Exergy Source (kW)	Exergy Sink (kW)	Exergy Loss (kW)	Exergy loss Percentage (%)
Furnace	36879.16	23611.04	13268.12	86.28
E7	2212.81	1641.36	571.45	3.72
E2	938.38	496.96	441.42	2.87
E13	1014.37	766.89	247.48	1.61
E1	505.37	341.98	163.40	1.06
E3	465.26	303.54	161.72	1.05
E6	335.45	193.55	141.91	0.92
E8	2136.32	2027.76	108.56	0.71
E10	909.99	814.73	95.25	0.62
E12	605.96	541.82	64.14	0.42
E4	102.39	59.47	42.92	0.28
E5	161.77	122.78	38.99	0.25
E11	270.46	251.36	19.10	0.12
E9	291.74	277.34	14.40	0.09
Total	46829.42	31450.57	15378.86	100

A Sankey diagram for CDU crude preheat train is illustrated in Figure 4.2. The exergy source is the exergy input and exergy sink is the exergy output of the system. Amount of exergy source is 46,829 kW and amount of exergy sink is 31,451 kW. The total exergy losses in CDU crude preheat train is 15,379 kW. The highest exergy loss in the system is furnace followed by heat exchanger E7, E2, E13 and the rest of heat exchangers.

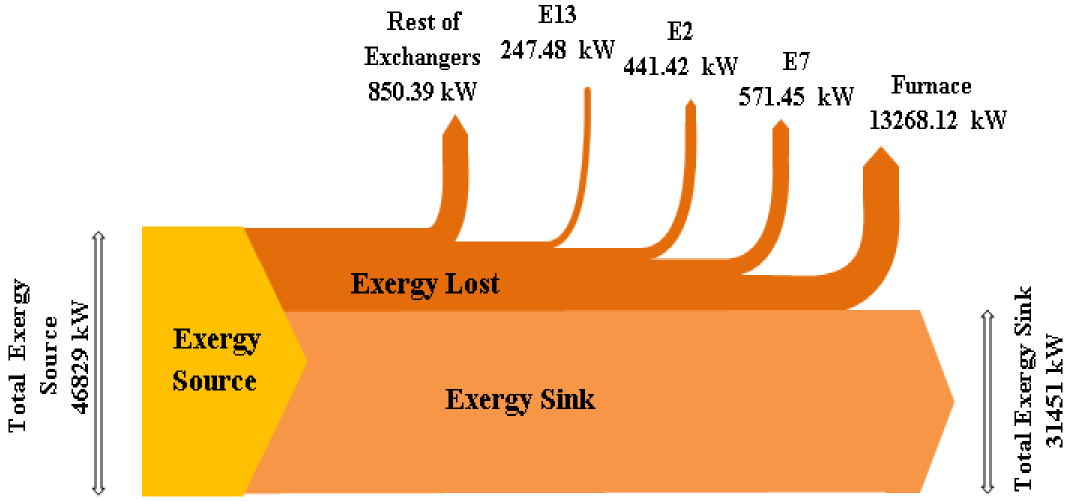

Figure 4.2 Sankey Diagram for CDU crude preheat train

4.1.3 Generation of Possible Fuel Reduction Strategies

The proposed method to reduce exergy loss of the system is to obtain closer gap between hot and cold composite curve. The justification of the options to minimize exergy loss is explained as follows.

a) Reducing Theoretical Flame Temperature (TFT)

Figure 4.3 shows the exergy composite curve where the hot composite curve is closer to the cold composite curve when TFT_o is reduced to TFT_i. Note that TFT_i is lesser than TFT_o. The strategy to reduce TFT is by using lower quality of fuel gas. However, reducing TFT is limited by the target temperature of tower inlet temperature (TIT). TFT should be hot enough to provide sufficient heat to the crude oil in order to meet the target temperature of TIT.

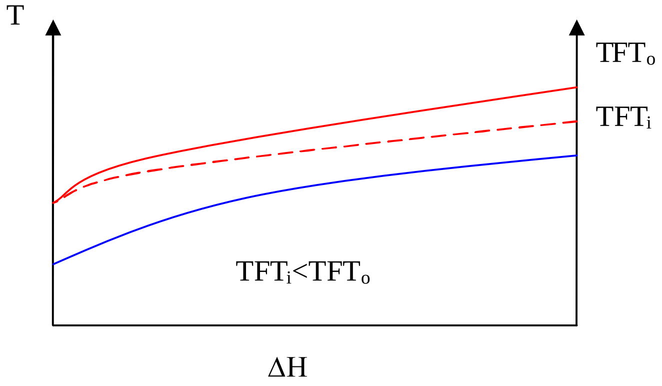

Figure 4.3 Hot composite curve is closer to cold composite curve when TFT_o is reduced to TFT_i

b) Reducing Stack Temperature (T stack)

Figure 4.4 shows the exergy composite curve where the hot composite curve is closer to the cold composite curve when T $stack_o$ is reduced to T $stack_i$. Note that T $stack_i$ is lesser than T $stack_o$. Usually the heat at the furnace stack is used as the heat source to preheat the boiling feed water for steam generation. This technique will reduce the stack temperature. However, reducing stack temperature is limited by the acid dew temperature of the flue gas. Stack temperature must always greater than acid dew temperature of flue gas to avoid environmental problem and corrosion.

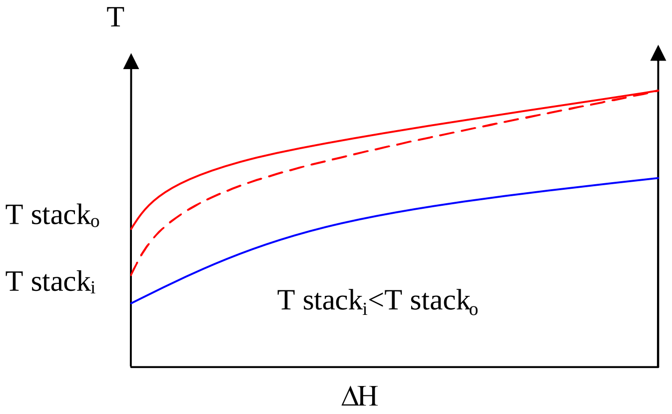

Figure 4.4 Hot composite curve is closer to cold composite curve when T stack$_o$ is reduced to T stack$_i$

c) Increasing Tower Inlet Temperature (TIT)

Figure 4.5 shows the exergy composite curve where the cold composite curve is closer to the hot composite curve when TIT$_o$ is increased to TIT$_i$. Note that TIT$_i$ is greater than TIT$_o$. The method to increase TIT is by increasing temperature difference of the furnace (ΔT). ΔT is the temperature difference between TIT and FIT. One way to increase ΔT is by reducing crude oil flow rate to the furnace. Heat duty is defined as $Q=mC_p\Delta T$ where flow rate (m) is inversely proportional to ΔT. Thus, decreasing flow rate (m) will increase ΔT of the furnace.

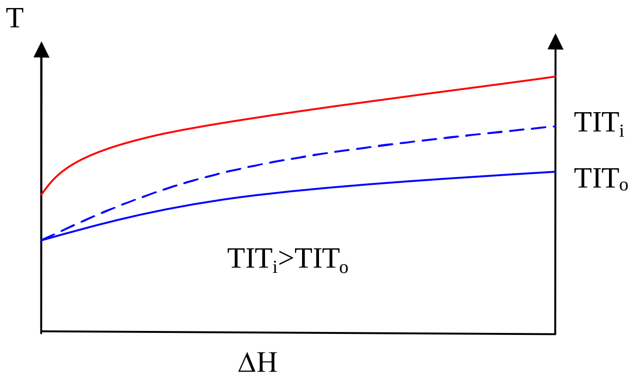

Figure 4.5 Cold composite curve is closer to hot composite curve when TIT$_o$ is increased to TIT$_i$

d) Increasing Furnace Inlet Temperature (FIT)

Figure 4.6 shows the exergy composite curve where the cold composite curve is closer to the hot composite curve when FIT_o is increased to FIT_i. Note that FIT_i is greater than FIT_o. The high heat transfer efficiencies of crude preheat train may increase FIT. The heat exchangers with high heat transfer efficiencies can be achieved by reducing the fouling rate. Cleaning heat exchangers at certain period of time may reduce the fouling rate.

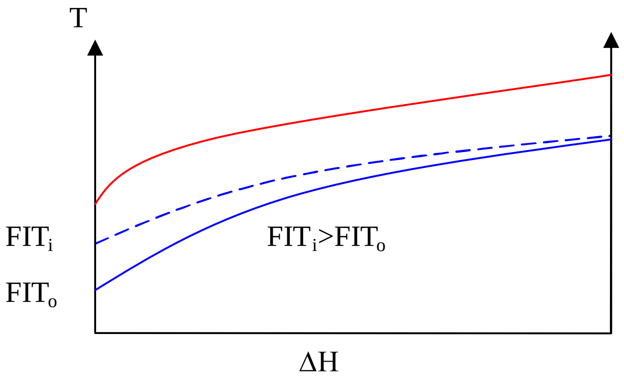

Figure 4.6 Cold composite curve is closer to hot composite curve when FIT_o is increased to FIT_i

4.1.3.1 Change in Exergy Loss for Possible Fuel Reduction Strategies

The change in exergy loss is calculated by plotting exergy loss against temperature. The slope of the graph is the change in exergy loss with unit of kW/^{o}C. The change in exergy loss is the quantitative measure of how much exergy is lost per degree Celsius of temperature. For example, the first fuel reduction strategy is reducing TFT as discussed before. The original value of TFT is 1615^{o}C. Six values of TFT with respective value of exergy losses are shown in Figure 4.7. The exergy losses for each value of TFT are obtained from Process Integration 'WORK' software.

From Figure 4.7, the change in exergy loss for reducing TFT option is 14.4 kW/^{o}C. If TFT is reduced by 1^{o}C, 14.4 kW of exergy loss is reduced.

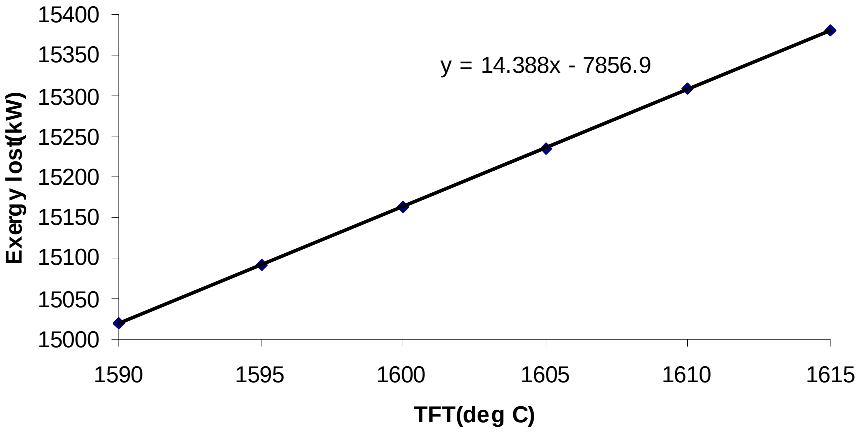

Figure 4.7 Exergy loss profile for reducing TFT

Table 4.5 summarized the amount of change in exergy loss per degree Celsius of temperature for all the possible fuel reduction strategies. The highest change in exergy loss is increasing FIT followed by increasing TIT, reducing TFT and reducing T stack. Thus, the possible fuel reduction options are able to reduce certain amount of exergy loss in order to achieve efficient operation.

Table 4.5 Comparison of the change in exergy loss for fuel reduction strategies

Options	Change in Exergy Loss (kW/^{o}C)
Reducing TFT	14.4
Reducing T stack	10.2
Increasing TIT	21.6
Increasing FIT	95.7

4.2 Fuel Reduction Strategies

4.2.1 Option 1: Reduction of Heat Loss from Furnace Stack

The flow rate of fuel gas, *m* is calculated using equation (3.13). Furnace stack temperature is plotted against volume flow rate of fuel gas. Figure 4.8 shows the relation between fuel gas flow rate and the furnace stack temperature. The slope of the graph is 2.6 $Nm^3/h\ ^oC$. This slope indicates that if stack temperature is reduced by 1^oC, 2.6 Nm^3/h of fuel gas can be saved.

The current stack temperature is 290^oC. If 50^oC reductions in stack temperature, the new stack temperature is 240^oC. Thus, the amount of fuel gas saving as 50^oC reduction in stack temperature is 130 Nm^3/h. In the plant historical data, the best practice for stack temperature is 169^oC. However, reducing stack temperature is limited by acid dew temperature of flue gas. Flue gas is the combustion products that leave the furnace through the furnace stack. The stack temperature must always higher than acid dew temperature of flue gas to avoid corrosion and environmental problems.

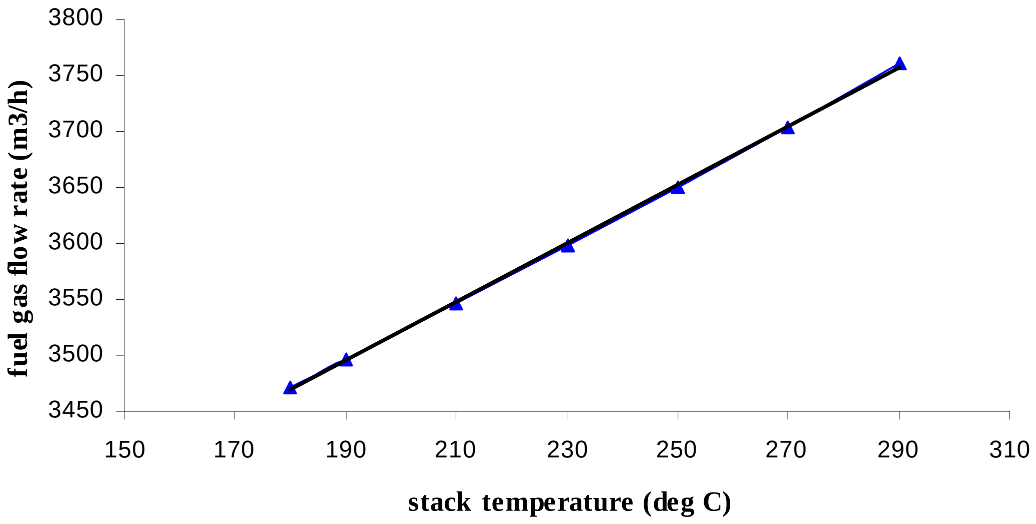

Figure 4.8 Fuel gas flow rate versus stack temperature of furnace

4.2.2 Option 2: Enhancement of Heat Recovery in the Crude Preheat Train

4.2.2.1 Case 1: Plant Shutdown for Crude Preheat Train Overall Cleaning

The model minimizes the expected cost throughout the time horizon. The cost is the trade off between extra fuel costs due to fouling, heat exchangers cleaning costs, loss in production costs due to plant shutdown and increased heat recovery after cleaning.

$$Cost = \alpha \sum_{t} (FG_t - FG_{t=0}) C_{fl} + \sum_{t} (y_t)(C_{cl} + C_{pl}) \tag{4.1}$$

FG_t (m^3/h) is the fuel gas consumption at time t, $FG_{t=0}$ is the fuel gas consumption at clean condition when period *t* equal to zero, C_{fl} (RM/GJ) is the furnace's fuel cost, C_{cl} is the cleaning cost and C_{pl} is the loss in production cost due to plant shutdown. Symbol α is the conversion factor with unit of GJ.h/month.m^3.

The binary variable y_t is defined to identify at what period *t* the crude preheat train is shutdown for cleaning.

$$y_t = \begin{cases} 1 & \text{if the plant is shutdown to clean crude preheat train in period } t \\ 0 & \text{otherwise} \end{cases}$$

The historical data for fuel gas flow rate and furnace inlet temperature (FIT) is collected. From the historical data, fuel gas flow rate is plotted against FIT to obtain quadratic correlation between these two variables. Fuel gas is expected to increase as FIT is decreased due to fouling in CPT as shown in Figure 4.9. Equation (4.2) shows the linear correlation between fuel gas flow rate and FIT.

$$FG_t = -31.997 \times (FIT_t) + 10139 \tag{4.2}$$

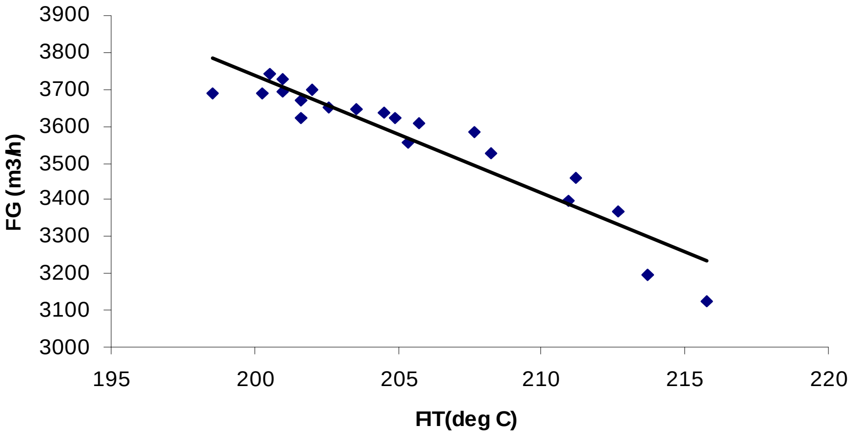

Figure 4.9 Linear correlation for fuel gas flow rate and FIT

Furnace inlet temperature (FIT) is the crude oil exit temperatures of crude preheat train before additional heating is provided by furnace. As time of operation increases, the value of FIT is expected to decrease due to the increment of fouling. FIT is reduced until it reaches threshold temperature. Threshold temperature is the critical temperature of chronic fouling condition. The value of critical temperature in this case is 200°C.

The graph of FIT versus month is plotted to obtain the slope of the graph. The slope is the amount of FIT reduction per month of operation as shown in Figure 4.10. The reduction of FIT per month is -0.56°C/month. Equation (4.3) is the reduction in FIT as time of operation increases.

$$FIT_t = (FIT_{t-1} - 0.5603)(1 - y_t) + FIT_{t=0}(y_t) \tag{4.3}$$

During any period when no cleaning is performed, y_t =0, the expression becomes

$$FIT_t = FIT_{t-1} - 0.5603 \tag{4.4}$$

Equation (4.4) indicates that FIT at period *t* is equal to FIT at previous period *t-1* deducted by 0.56.

When cleaning is performed at certain period *t*, y_t = 1, the expression reduces to

$$FIT_t = FIT_{t=0} \tag{4.5}$$

Equation (4.5) shows that FIT at period t is equal to FIT at period t=0 when cleaning is performed. FIT at t=0 is the maximum value of FIT at clean condition.

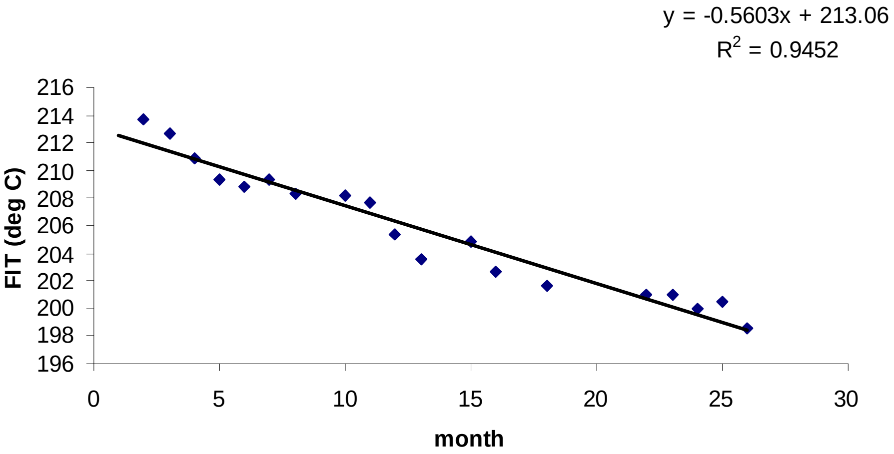

Figure 4.10 Linear trends for FIT profile

The initial assigned value of FIT at t=0 is the maximum value of FIT in the refinery at clean condition which is at 215°C.

$$FIT_{t=0} = 215 \tag{4.6}$$

The minimum allowable FIT is greater than 200°C. FIT at any period of *t* should not less than critical temperature that is determined by plant engineer in the refinery.

$$FIT_t \geq 200 \tag{4.7}$$

4.2.2.2 Case 2: Heat Exchangers Online Cleaning

The objective function minimizes cost that includes furnace's extra fuel cost and cleaning cost. Equation (4.8) represents the objective function for online cleaning schedule.

$$Cost = \alpha \sum_t (FG_t - FG_{t=0}) C_{fl} + \sum_t \sum_i y^{cl}_{i,t} C_{cl} \tag{4.8}$$

The binary variable y_{it}^{cl} is defined to identify when and which heat exchanger is cleaned.

$$y_{i,t}^{cl} = \begin{cases} 1 & \text{if the } i^{th} \text{ heat exchanger is cleaned in period } t \\ 0 & \text{otherwise} \end{cases}$$

The fouling behaviour of individual units in the preheat train is investigated by analyzing operating data collected over 10 months.

$$Q = UA\Delta T_{lm} \tag{4.9}$$

$$\frac{1}{Uf_{i,t}} = \frac{1}{Uc_i} + Rf_{i,t} \tag{4.10}$$

The values of the clean heat transfer coefficient, Uc_i are calculated using equation (4.9) where all heat exchangers are in clean condition. The historical data for cleaned heat exchangers are collected in the first month of plant operation after refinery's turnaround. The fouled heat transfer coefficient, Uf_{it} is calculated using equation (4.10). Fouling resistance value for all heat exchangers in CPT for 10 months operation is shown in Appendix A.

Figure 4.8 shows fouling resistance profiles for heat exchangers E1 until E13. The trend of the profiles suggests that linear fouling is occurred. The slope of the profile indicates the rate of fouling per month. The fouling rates are summarized in Table 4.6. Equation (4.11) is the linear fouling resistance equation:

$$Rf_{i,t} = (Rf_{i,t-1} + dRf_{i,t})(1 - y_{i,t}^{cl}) + (Rf_{i,t=0})(y_{i,t}^{cl}) \tag{4.11}$$

$Rf_{i,t}$ it is the fouling resistance for i^{th} heat exchanger in period t, $Rf_{i,t-1}$ is the fouling resistance in previous period t. The slope of graph in Figure 4.11 gives the value for fouling rate, $dRf_{i,t}$. Fouling resistance in period t=0 is $Rf_{i,0}$.

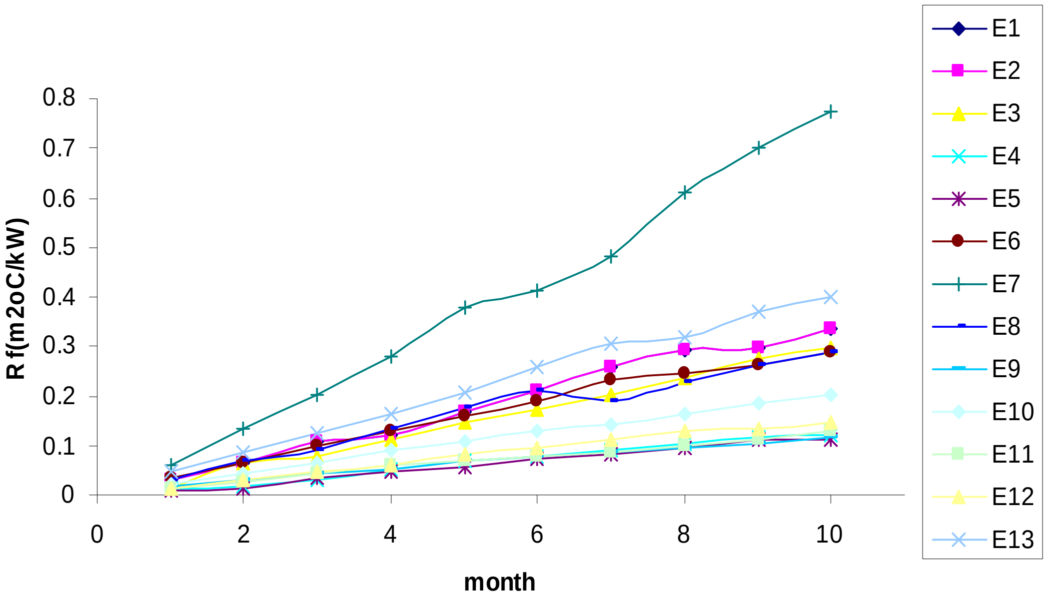

Figure 4.11 fouling resistance profiles for crude preheat train

The initial fouling resistance, $Rf_{i,0}$ for all i^{th} heat exchangers is set to zero. Thus, at initial period $t=0$, $y^{cl}_{i,t=0}$ must equal to 1.

When, $y^{cl}_{i,t} = 1$, the expression reduced to

$$Rf_{i,t} = Rf_{i,0} \tag{4.12}$$

The i^{th} heat exchanger is being cleaned at period t, thus $Rf_{i,t}$ equals to $Rf_{i,0}$.

The correlation between fuel gas flow rate and FIT is the same equation as in Case 1. FIT is the outlet cold stream temperature (Tc2) for heat exchanger E13. The respective equations are as follows:

$$FG_t = -31.997 \times (Tc2_{E13,t})^2 + 10139 \tag{4.13}$$

$$FIT_t = (Tc2_{E13,t-1} - 0.5603)(1 - y^{cl}_{i,t}) + Tc2_{E13,t=0}(y^{cl}_{i,t}) \tag{4.14}$$

Table 4.6 Fouling data for crude preheat train

Heat exchanger	Heat exchanger type	Area, A_i (m^2)	Clean heat transfer coefficient, Uc_i ($kWm^2/^oC$)	Fouling rate, dRf_{it} ($m^{2o}C/kW$)
E1	Shell and tubes	411.5	1.073	0.035
E2	Shell and tubes	111.0	0.660	0.060
E3	Shell and tubes	361.8	0.095	0.033
E4	Shell and tubes	111.0	0.033	0.013
E5	Shell and tubes	180.9	0.100	0.012
E6	Shell and tubes	111.4	0.085	0.032
E7	Shell and tubes	132.5	0.736	0.070
E8	Shell and tubes	125.0	4.470	0.030
E9	Shell and tubes	111.4	0.542	0.011
E10	Shell and tubes	169.9	0.320	0.020
E11	Shell and tubes	153.7	0.185	0.012
E12	Shell and tubes	165.5	0.212	0.015
E13	Shell and tubes	78.2	0.227	0.040

The equations that represent the relationship between the inlet and the outlet temperatures of the ith heat exchanger are the heat duty as in equation (4.15) and (4.16).

$$Q_{i,t} = Fc_i Cc_i (Tc2_{i,t} - Tc1_{i,t}) \tag{4.15}$$

$$Q_{i,t} = U_{i,t} A_i \frac{(Th1_{i,t} - Tc2_{i,t}) - (Th2_{i,t} - Tc1_{i,t})}{\ln\left[\dfrac{(Th1_{i,t} - Tc2_{i,t})}{(Th2_{i,t} - Tc1_{i,t})}\right]} \tag{4.16}$$

These equations are rearranged to get the equations for outlet hot stream temperature, $Th2_{,i,t}$ and outlet cold stream temperature, $Tc2_{,i,t}$. Equation (4.17) and (4.18) are obtained from Lavaja & Bagajewicz (2004).

$$Th2_{i,t} = \frac{(R-1)Th1_{,i,t} + \left\{\exp\left[\dfrac{Uf_{i,t}A_i}{Fc_i Cc_i}(R-1)\right] - 1\right\} RTc1_{,i,t}}{R \times \exp\left[\dfrac{Uf_{i,t}A_i}{Fc_i Cc_i}(R-1)\right] - 1} \tag{4.17}$$

where $$R = \frac{Fc_i Cc_i}{Fh_i Ch_i} = \frac{Th1_{i,t} - Th2_{i,t}}{Tc2_{i,t} - Tc1_{i,t}}$$

$$Tc2_{i,t} = Tc1_{i,t} + \frac{Th1_{i,t} - Th2_{i,t}}{R} \tag{4.18}$$

There are four inequality constraints in this scheduling formulation. The first constraint is FIT must always be greater than minimum allowable FIT at 200°C. FIT is the outlet cold stream temperature for heat exchanger E13.

$$Tc2_{E13,t} \geq 200 \tag{4.19}$$

The second constraint is the maximum fouling resistance for parallel heat exchangers. Equation (4.20) demonstrates that only parallel heat exchangers which are heat exchanger E4 and E8 have maximum limit of allowable fouling resistance because only these two heat exchangers are allowed to perform online cleaning. While, the rest of heat exchangers are not considered to be cleaned. The maximum fouling rate is obtained from manual calculation when FIT reached 200°C.

$$\begin{aligned} Rf_{E4,t} &\leq 0.22 \\ Rf_{E8,t} &\leq 0.51 \end{aligned} \tag{4.20}$$

Table 4.7 shows the maximum fouling resistance for all heat exchangers when FIT is 200°C.

Table 4.7 Maximum fouling resistance for heat exchanger

Heat exchanger	Maximum fouling resistance, Rfit ($m^{2o}C/kW$)
E1	0.60
E2	1.02
E3	0.56
E4	0.22
E5	0.20
E6	0.54
E7	1.19
E8	0.51
E9	0.19
E10	0.34
E11	0.20
E12	0.26
E13	0.68

The third constraint is only heat exchanger E4 and E8 is allowed to perform online cleaning because only these two heat exchangers are parallel heat exchangers. The maximum number of cleaning for heat exchanger E4 and E8 is 60 which equivalent to 60 months or 5 years of operations. Equation (4.21) provides the upper limit of the total number of cleaning for heat exchanger E4 and E8.

$$\begin{aligned} \sum_t y^{cl}_{i=E4,t} &\leq 60 \\ \sum_t y^{cl}_{i=E8,t} &\leq 60 \end{aligned} \tag{4.21}$$

Other heat exchangers are not allowed to be cleaned. Equation (4.22) shows that total number of cleaning for other heat exchangers except for heat exchanger E4 and E8 are less or equal to 1.

$$\sum_t y^{cl}_{i \neq E4 \& E8,t} \leq 1 \tag{4.22}$$

4.2.2.3 Case 3: Heat Exchangers Online Cleaning with Retrofit of High Efficiency Heat Exchangers

The purpose of scheduling formulation for Case 3 is to identify which heat exchanger is selected for retrofit with high efficiency heat exchanger while performing online cleaning throughout the time horizon. The objective function is minimizing operational cost which is the same as the objective function in case 2. The differences are found in several equations to formulate the changes of previous conventional shell and tubes exchangers into new high efficiency heat exchangers such as Compabloc welded plate heat exchangers.

$$Cost = \alpha \sum_t (FG_t - FG_{t=0}) C_{fl} + \sum_t \sum_i y_{i,t}^{cl} C_{cl} \tag{4.23}$$

Equation (4.13), (4.14), (4.17) and (4.18) are used in the formulation for Case 3. These equations are correlation between fuel gas flow rate and FIT, FIT reduction equation, outlet hot stream temperature equation and outlet cold stream temperature equation.

The additional term in overall heat transfer coefficient equation is the substitution of binary variable, $y_{i,t}^{cg}$. The second binary variable, $y_{i,t=0}^{cg}$ in case 3 is to determine which heat exchanger is selected to change into high efficiency heat exchanger in period t =0. The purpose of changing high efficiency heat exchanger in period t =0 is to formulate new online cleaning after the installation of high efficiency heat exchangers at initial period.

$$y_{i,t=0}^{cg} = \begin{cases} 1 & \text{if the } i^{th} \text{ heat exchanger is changed in period } t = 0 \\ 0 & \text{otherwise} \end{cases}$$

Equation (4.24) as shown below is due to the assumption that clean heat transfer coefficient for high efficiency heat exchanger is twice greater than previous clean heat transfer coefficient for shell and tubes heat exchanger (Arvidsson P., 2003).

$$Uf_{i,t} = \frac{1}{\frac{1}{Uc_i + Uc_i \times \sum_t y^{cg}_{i,t}} + Rf_{i,t}} \tag{4.24}$$

For example, if summation of $y^{cg}_{i,t=0}$ is equal to 1. Then $Uf_{i,t}$ is:

$$Uf_{i,t} = \frac{1}{\frac{1}{Uc_i + Uc_i} + Rf_{i,t}} \tag{4.25}$$

The application of typical fouling resistance, *Rf* for conventional shell and tube exchangers are recommended by Tubular Exchanger Manufacturers Association (TEMA). However, Hesselgreaves (2002) recommended the fouling resistance value for compact heat exchangers, such as plate and frame heat exchanger, are in the order of one tenth of TEMA values. Thus, when $y^{cg}_{it=0}$ is selected, Rf_{it} is reduced by one tenth or 10% from the previous Rf_{it}.

The term $dRfhex_i$ in equation (4.26) is the value of 90% reduction of fouling rates for i^{th} high efficiency heat exchanger.

$$dRfhhx_{i,t} = dRfhex_i \times y^{cg}_{i,t=0} \tag{4.26}$$

The additional term for fouling resistance equation is $dRfhhx_{i,t}$. This is the term added in equation (4.27) to indicate if $y^{cg}_{i,t=0}$ is selected or not selected.

$$Rf_{i,t} = (Rf_{i,t-1} + dRf_{i,t} - dRfhhx_{i,t})(1 - y^{cl}_{i,t})(1 - y^{cg}_{i,t}) + (Rf_{i,0})(y^{cl}_{i,t} + y^{cg}_{i,t}) \tag{4.27}$$

If $y^{cg}_{i,t=0}$ is not selected, ($y^{cg}_{i,t=0}$ =0) then, equation (4.26) and (4.27) are expressed as below.

$$dRfhhx_{i,t} = 0 \tag{4.28}$$

$$Rf_{i,t} = (Rf_{i,t-1} + dRf_{i,t})(1 - y^{cl}_{i,t}) + (Rf_{i,0})(y^{cl}_{i,t}) \tag{4.29}$$

The value of fouling rates, $dRf_{i,t}$ for i^{th} heat exchanger and value of 90% reduction in fouling rates, $dRfhex_{i,t}$ for i^{th} high efficiency heat exchanger is shown in Table 4.8.

Table 4.8 Value of fouling rates for previous shell and tubes and value of reduction in fouling rates for high efficiency heat exchanger

Heat exchanger	dRf_{it} ($m^{2o}C/kW$)	$dRfhex_i$ ($m^{2o}C/kW$)
E1	0.035	0.0315
E2	0.06	0.054
E3	0.033	0.0297
E4	0.013	0.0117
E5	0.012	0.0108
E6	0.032	0.0288
E7	0.07	0.063
E8	0.03	0.027
E9	0.011	0.0099
E10	0.02	0.018
E11	0.012	0.0108
E12	0.015	0.0135
E13	0.04	0.036

The inequality constraints in this formulation are as follows:

FIT should be greater than minimum allowable FIT at 200°C. This value of minimum allowable FIT is decided by plant engineer in refinery.

$$Tc2_{E13,t} \geq 200 \tag{4.30}$$

Either cleaning i^{th} heat exchanger, y_{it}^{cl} or changing i^{th} high efficiency heat exchanger, $y_{i,t}^{cg}$ is selected in any period t. Equation (4.31) is inserted in the formulation to avoid both cleaning i^{th} heat exchanger and changing i^{th} high efficiency heat exchanger are selected in the same period t.

$$y_{i,t}^{cl} + y_{i,t}^{cg} \leq 1 \tag{4.31}$$

Equation (4.32) shows that changing i^{th} high efficiency heat exchanger binary variable, $y_{i,t}^{cg}$ is allowed in period t=0 only.

$$y_{i,t=0}^{cg} \leq 1 \tag{4.32}$$

Changing i^{th} high efficiency heat exchanger binary variable, y_{it}^{cg} for all period t except at initial period is not allowed. Equation (4.33) demonstrates that changing i^{th} high efficiency heat exchanger is not allowed at all period t except at initial period. The i^{th} high efficiency heat exchanger is only allowed to change at initial period as shown in equation (4.32)

$$\sum_t y_{i,t+1}^{cg} \leq 0 \tag{4.33}$$

All heat exchangers are allowed to clean or change into high efficiency heat exchanger only once for all period t except for parallel heat exchangers E4 and E8.

$$\sum_t (y_{i \neq E4\&E8,t}^{cl} + y_{i \neq E4\&E8,t}^{cg}) \leq 1 \tag{4.34}$$

4.2.3 Carbon Credit

The carbon credit or carbon emission reduction (CER) revenue is calculated for all the cases. CERs are carbon credits issued by the Clean Development Mechanism (CDM) Executive Board for emission reductions achieved by CDM projects and verified by a DOE under the rules of the Kyoto Protocol. The equation to calculate CER revenue is as below.

$$(CO2_{current\ practice} - CO2_{options})R_{CER} \tag{4.35}$$

$CO2_{current\ \mathrm{practice}}$ is the carbon dioxide emission for current practice. $CO2_{options}$ is the carbon dioxide for all the options and R_{CER} is the revenue of certified emission reduction. The value of CER in 2009 is €13/tonne or approximately RM 55/tonne as reported in Point Carbon 5th annual conference (Tvinnereim et al., 2009).

4.3 Feasibility Analysis for Fuel Reduction Strategies

The current practice in the refinery to reduce energy consumption due to fouling is to shutdown the plant for crude preheat train (CPT) overall cleaning. The fouling rates formation in the crude preheat train are observed by monitoring furnace inlet temperatures (FIT). When FIT reaches the minimum allowable temperature which is below 200°C, the engineers may consider to shutdown the plant for crude preheat train cleaning. The current practice to minimize energy consumption is non-optimal overall cleaning schedule for CPT because the engineers do not use any optimization tool to predict when the optimal period to shutdown the plant for CPT cleaning. The refinery's turnaround is every five years. For the last five years, the plant was shutdown twice to clean CPT. Table 4.11 shows the overall cleaning schedule for current practice.

In feasibility analysis, comparison of energy and carbon dioxide emission saving are made between all the options with current practice to verify that the fuel reduction options are able to reduce energy consumption and carbon dioxide emission in the CDU. The profiles of FIT and extra fuel gas flow rate (ΔFG) for all the options are shown to observe the trends of the profiles.

4.3.1 Energy and Carbon Dioxide Emission Saving

Table 4.9 demonstrates the comparison of energy saving for all options. Extra fuel gas flow rate (ΔFG) is the total amount of additional fuel gas needed due to fouling in crude preheat train. The amount of energy is calculated by multiplying extra fuel gas flow rate (Nm^3/year) with heating value of fuel gas. The heating value is 14.95 kWh/kg or 0.029 GJ/Nm^3 as shown in Table 3.6. Energy saving is the difference between energy of current practice and energy of the options.

For option 1, 1°C reduction in stack temperature will save 2.6 Nm^3/h of fuel gas as explained in Figure 4.8. The amount of fuel gas saving as 50°C reduction in stack temperature is 130 Nm^3/h which equal to energy saving of 45,614 GJ/year. Therefore, the percentage of energy saving for Option 1 is 1.12%.

For Option 2, the amounts of extra fuel gas flow rate are obtained from optimization model. The list of fuel gas flow rate at period t=0 until t=60 for all cases in option 2 are shown in Appendix C. The percentage of energy saving for Case 1, Case 2 and Case 3 are 6%, 14% and 60%, respectively.

Table 4.10 shows the comparison of carbon dioxide emission saving for all options. The carbon dioxide extra flow rate is calculated by using equation (3.15). Carbon dioxide emission saving per year is the difference between carbon dioxide flow rate of current practice and carbon dioxide flow rate of the options.

From the comparison, Case 3 in Option 2 is the largest percentage of energy and carbon dioxide emission savings followed by Case 2, Case 1 and Option 1. The amount of extra fuel gas and carbon dioxide emission in Case 3 are greatly reduced compared to other options. The installation of high efficiency heat exchangers in Case 3 improves furnace inlet temperature (FIT) by having lower fouling rates and results to the highest energy saving and the lowest carbon dioxide emission.

Table 4.9 Comparison of energy saving for all options

Options	Fuel gas			
	Extra flow rate (Nm^3/year)	Amount of Energy (GJ/year)	Energy saving (GJ/year)	Percentage saving (%)
Current Practice	99,598,913	2,888,368	-	-
Option 1	98,478,720	2,855,883	32,486	1.12
Option 2: Case 1	94,022,631	2,726,656	161,712	5.60
Option 2: Case 2	85,263,468	2,472,641	415,728	14.39
Option 2: Case 3	40,009,715	1,160,282	1,728,087	59.83

Table 4.10 Comparison of carbon dioxide emission saving for all options

Options	Carbon dioxide		
	Extra flow rate (kg/year)	Saving per year (kg/year)	Percentage saving (%)
Current Practice	187,013,808	-	-
Option 1	184,910,455	2,103,352	1.12
Option 2: Case 1	176,543,395	10,470,412	5.60
Option 2: Case 2	160,096,585	26,917,222	14.39
Option 2: Case 3	52,624,818	78,377,732	59.83

4.3.2 FIT Profiles

Figure 4.12 demonstrates FIT profiles for current practice and Case 1. FIT decreases gradually as period *t* increases due to fouling in crude preheat train. However, FIT is increased to its maximum value during cleaning period *t*. Based on plant historical data, the maximum value of FIT at clean condition is 215^{o}C. The initial period at *t* = 0 and the final period at *t* = 60 are the period of operation at clean condition. From Figure 4.12, FIT is increased to 215°C at 17th months and 44th months for Case 1 and at 28th months and 49th months for current practice.

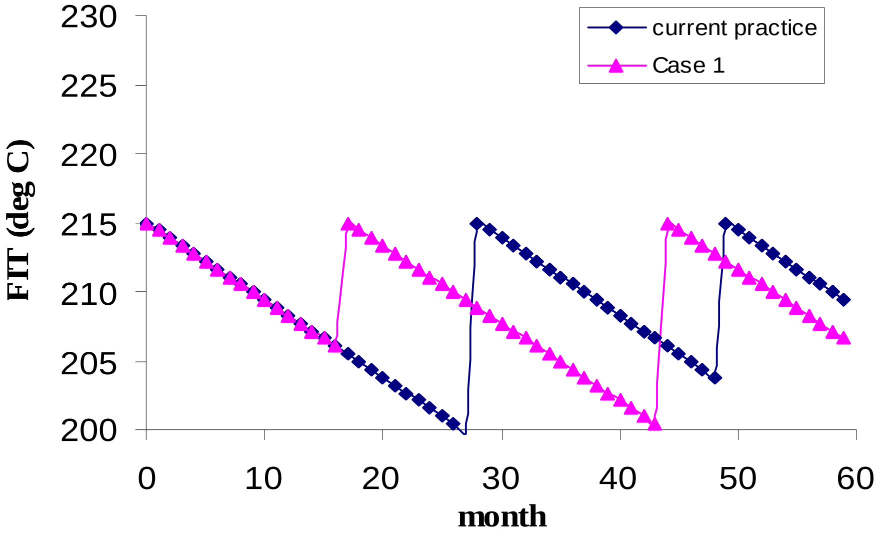

Figure 4.12 FIT profiles for Current Practice and Case 1

Figure 4.13 shows FIT profiles for Case 2 and Case 3. During operations, FIT decreases linearly. After performing online cleaning, FIT increases slightly.

From Figure 4.13, value of FIT for Case 3 at initial period is higher than FIT for Case 2. This is due to the installation of high efficiency heat exchangers at the beginning of plant operation. The installation of new high efficiency heat exchangers improves FIT from 215°C to 227°C due to higher overall heat transfer coefficient and lower fouling rates than shell and tubes heat exchangers.

For the overall plant performance, FIT profile for Case 3 is smaller decay than FIT profile for Case 2. At the end of operation, FIT is relatively high at 222°C. Throughout the operation, temperature drop of FIT for Case 2 is 10°C while temperature drop for Case 3 is only 5°C. List of FIT values for all cases are shown in Appendix C.

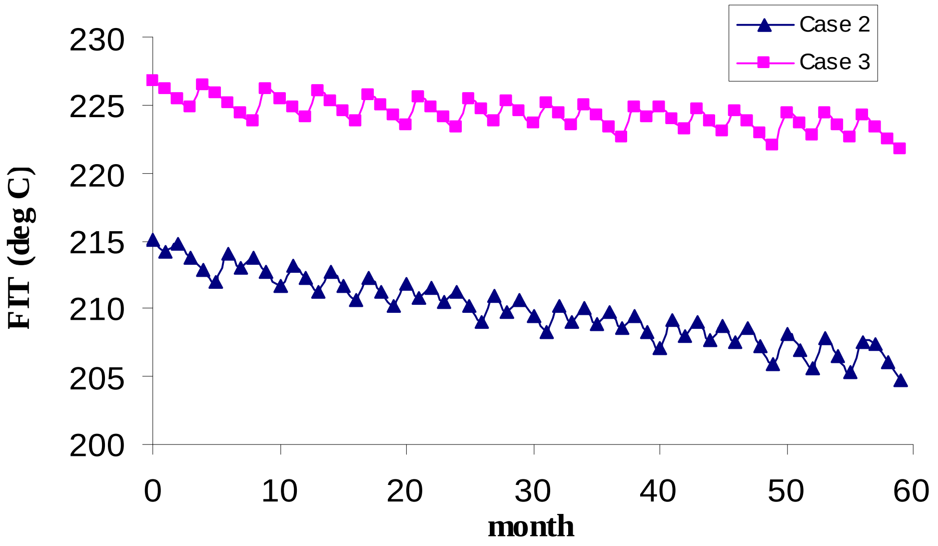

Figure 4.13 FIT profiles for Case 2 and Case 3

4.3.3 Extra Fuel Gas Profiles

Figure 4.14 shows extra fuel gas profiles for current practice and Case 1. Extra fuel gas (ΔFG) is increasing throughout the operation. There is no extra fuel gas after cleaning period which is at 17th and 44th months for Case 1 and 28th and 49th months for current practice. The total amount of extra fuel gas throughout the operation is 11,528 Nm^3/h for current practice and 10,882 Nm^3/h for Case 1.

Figure 4.15 shows extra fuel gas profiles for Case 2 and Case 3. Total amount of extra fuel gas throughout the operation is 9,868 Nm^3/h for Case 2 and 4,631 Nm^3/h for Case 3. The amount of extra fuel gas for Case 3 is greatly reduced because of the installation of high efficiency heat exchangers. Extra fuel gas for Case 2 is slightly less than Case 1 because only parallel heat exchangers are performed online cleaning while the rest of heat exchangers severe greater fouling rates. Furthermore, in Case 1, the reduction of a few of extra fuel gas is due to earlier crude preheat train overall cleaning schedule than current practice.

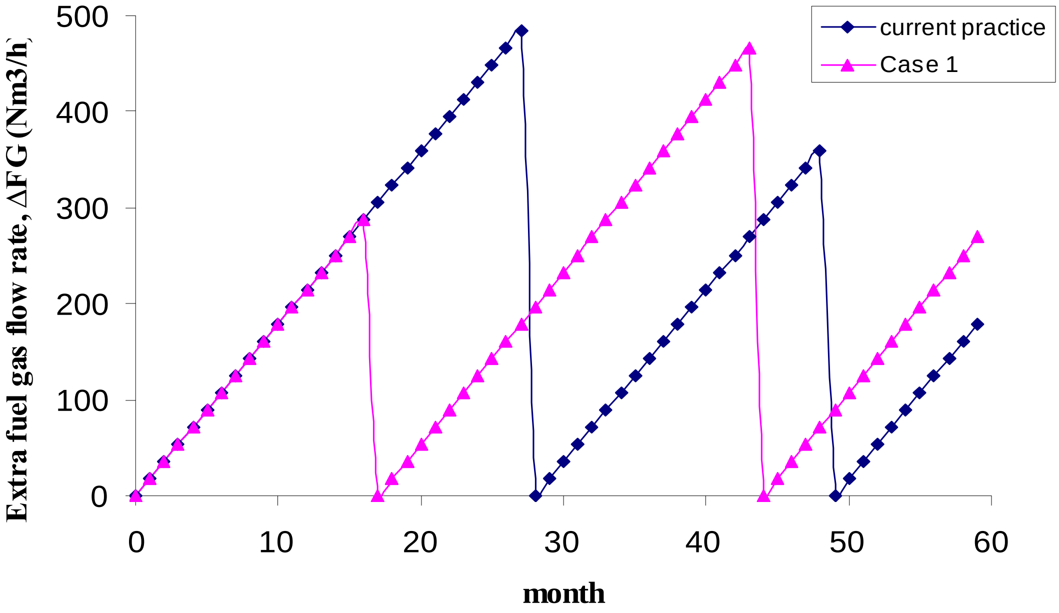

Figure 4.14 Extra fuel gas profiles for current practice and Case 1

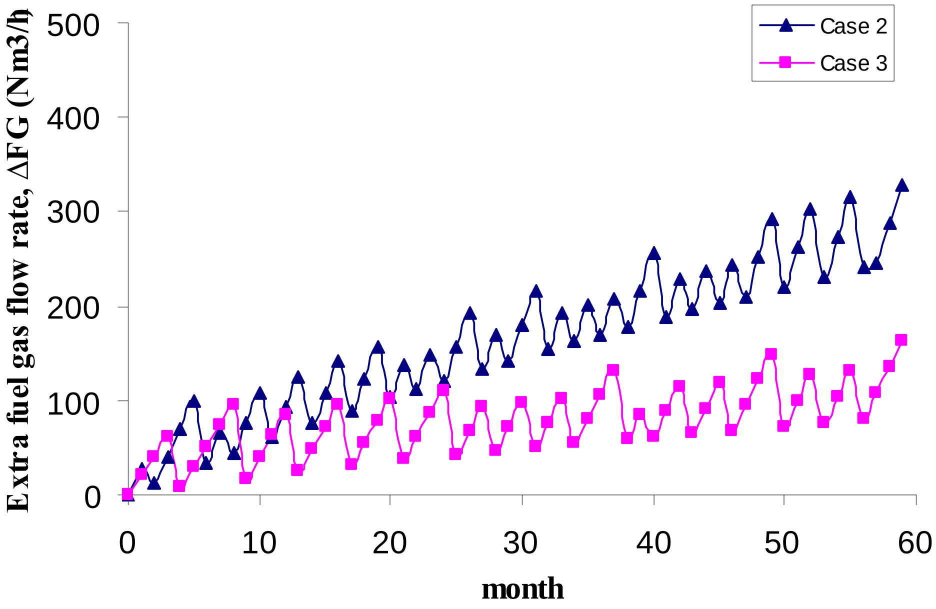

Figure 4.15 Extra fuel gas profiles for Case 2 and Case 3

4.3.4 Heat Exchanger Cleaning Schedules

From Table 4.11, cleaning schedule for Case 1 is on 17th and 44th months. Meanwhile for current practice, the cleaning schedule is on 28th (August 2007) and 49th (May 2009) months.

The purpose of only performing online cleaning for heat exchangers in parallel position is to maintain throughput flow rate of the plant. In parallel position, one of the heat exchanger is in operation mode while the other one is on standby mode. When online cleaning is performed, the parallel heat exchanger on standby mode is changed to operation mode. Table 4.12 shows the optimal cleaning schedule for Case 2. Only heat exchangers in parallel position are allowed for cleaning. These are heat exchanger E4 and E8. There are 27 numbers of cleaning in the schedule for 60 months period. Fouling rates, dRf for heat exchanger E8 is higher than heat exchanger E4 as shown in Table 4.7. Therefore, heat exchanger E8 has higher increment in fouling resistance, Rf which will significantly reduces FIT over time. As a result, E8 is cleaned regularly compared to E4 to minimize operational cost and maintaining FIT above 200°C.

The optimal online cleaning schedule for Case 3 is shown in Table 4.13. Total number of cleaning is 16. Only heat exchanger E8 is cleaned regularly due to high fouling rates. Nine heat exchangers are selected to change into high efficiency heat exchangers. The selected heat exchangers are E1, E4, E5, E6, E9, E10, E11, E12, and E13. Heat exchanger E4, which is heat exchanger in parallel position, is also selected. The selected heat exchangers are capable to improve plant performance and reduce operational cost.

Table 4.11 Current practice and optimal cleaning schedule for Case 1

	month									No. of cleaning
	1-16	17	18-27	28	29-43	44	45-48	49	50-59	
Current Practice				•				•		2
Case 1		•				•				2

Table 4.12 Optimal cleaning schedule for Case 2

Hex	month																											No. of cleaning
	2	6	8	11	12	14	17	20	22	24	26	27	29	32	34	36	37	38	41	43	45	47	49	50	53	56	57	
E1																												0
E2																												0
E3																												0
E4					•						•						•						•					4
E5																												0
E6																												0
E7																												0
E8	•	•	•	•		•	•	•	•	•		•	•	•	•	•		•	•	•	•	•		•	•	•	•	23
E9																												0
E10																												0
E11																												0
E12																												0
E13																												0
																											Total No. of Cleaning	27

Table 4.13 Optimal cleaning schedule for Case 3

Hex	month																No. of cleaning
	4	9	13	17	21	25	28	31	34	38	40	43	46	50	53	56	
E1																	0
E2																	0
E3																	0
E4																	0
E5																	0
E6																	0
E7																	0
E8	•	•	•	•	•	•	•	•	•	•	•	•	•	•	•	•	16
E9																	0
E10																	0
E11																	0
E12																	0
E13																	0
														Total No. of Cleaning			16

4.4 Economic Analysis for Fuel Reduction Strategies

The current practice in the refinery is to shutdown the plant for overall cleaning of CPT. For the last five years, the plant is shutdown twice to clean CPT. The total cost calculation for current practice is the summation of fuel gas cost, heat exchanger cleaning cost, lost in production cost due to fouling and opportunity loss cost due to plant shutdown.

The current practice cost is calculated as demonstrated in Table 4.14. In conversion column, the value of 0.029 GJ/m^3 is multiplied by 24 hour/day and 30 day/month. From the plant historical data on fuel gas price, cleaning cost and production lost cost, the value of fuel gas price is RM 14.55/GJ, cleaning cost is RM 40,000/unit and production lost cost is RM 545,000/day. CDU is expected to shutdown for 10 days during crude preheat train cleaning. The annual total cost for current practice is RM 90,254,588/year.

Table 4.14 Current practice cost calculation

details	amount	cost	multiply	conversion	RM
Total extra fuel gas flow rate	11,528 Nm^3/h	RM 14.55/ GJ	60 months	20.88 GJ.h/month.m^3	210,135,151
Heat exchanger cleaning	13 units	RM 40,000/unit	2 times	-	1,040,000
Production lost	10 days	RM 545,000/day	2 times	-	10,900,000
		Total (RM)			222,075,151
		Annual cost (RM/year)			44,415,030

Table 4.15 and Table 4.16 demonstrate the cost for all options without and with carbon credit. The comparison shows that Case 3 results in the highest percentage of cost saving which are about 71% and 62% with and without carbon credit, respectively. The next cost saving options is Case 2, Case 1 and Option 1. Percentage of cost saving with carbon credit shows higher percentage than without carbon credit.

For Option 2, it is expected that Case 1 is the lowest cost percentage saving. The earlier cleaning schedule than current practice cleaning schedule causes little reduction in extra fuel gas flow rate. The cost percentage saving for Case 2 is higher than Case 1 because frequent online cleaning for parallel heat exchangers is able to delay the reduction of FIT due to fouling and thus reducing extra fuel gas flow rate. However, cost percentage saving for Case 2 is much smaller than Case 3. The great amount of extra fuel gas is reduced due to the usage of high efficiency heat exchangers with very low fouling rates in Case 3. In addition, online cleaning for Case 2 is only limited to two parallel heat exchangers but the other heat exchangers experienced severe increment in fouling as time of operation increases.

Table 4.15 Cost for all options without Carbon Credit

Options	Total cost (RM)	Annual cost (RM/year)	Cost Saving per year (RM/year)	Percentage of cost saving (%)
Current Practice	222,075,151	44,415,030	-	-
Option 1	207,765,480	41,553,096	2,861,934	6.44
Option 2: Case 1	210,290,613	42,058,123	2,356,908	5.31
Option 2: Case 2	180,952,239	36,190,448	8,224,582	18.52
Option 2: Case 3	85,244,694	17,048,939	27,366,091	61.61

Table 4.16 Cost for all options with Carbon Credit

Options	CO_2 revenue (RM)	Total cost (RM)	Annual cost (RM/year)	Cost Saving per year (RM/year)	Percentage of cost saving (%)
Current Practice	-	222,075,151	44,415,030	-	-
Option 1	578,422	207,187,058	41,437,412	2,977,619	6.70
Option 2: Case 1	2,879,363	207,411,250	41,482,250	2,932,780	6.60
Option 2: Case 2	7,402,236	173,550,003	34,710,001	9,705,030	21.85
Option 2: Case 3	21,553,876	63,690,817	12,738,163	31,676,867	71.32

The purchase cost for high efficiency heat exchangers are calculated to obtain payback period. In refinery's crude preheats train, the most common high efficiency heat exchangers is Alfa Laval Compabloc welded plate heat exchanger.

Table 4.17 summarizes the purchase cost for selected heat exchangers in Case 3. The purchase cost equation for Compabloc heat exchanger is shown in equation (3.17). The total investment cost for purchasing Compabloc welded plate heat exchangers are RM 8,903,512. The payback period calculation is total investment cost over the saving cost. The saving cost for Case 3 without and with carbon credit are

RM 27,366,091/year and RM 31,676,867/year, respectively. As a result, the payback period without carbon credit is 4 months and the payback period with carbon credit is 3 months.

Table 4.17 Purchase cost calculation for Compabloc welded plate heat exchangers

Heat exchanger	A (m^2)	A (ft^2)	Cp ($)
E1	411.5	4,429.221	434,721
E4A	111	1,194.76	228,235
E4B	111	1,194.76	228,235
E5	180.9	1,947.135	290,193
E6	111.4	1,199.065	228,636
E9	111.4	1,199.065	228,636
E10	169.9	1,828.736	281,377
E11	153.7	1,654.365	267,847
E12	165.5	1,781.376	277,770
E13	78.2	841.7135	192,120
Total Purchase cost ($)			2,657,765
Total purchase cost (RM)			8,903,512

CHAPTER 5

CONCLUSION AND FUTURE WORKS

5.1 Conclusion

In summary, the most cost-effective and most practical option for fuel reduction strategies in crude distillation unit is enhancement of heat recovery in the crude preheat train (CPT) by optimal heat exchangers online cleaning schedule with retrofit of high efficiency heat exchangers (Case 3). Case 3 is the highest percentage of energy and carbon dioxide emission savings as well as the lowest percentage of operational cost. This is due to the replacement of selected conventional shell and tube heat exchangers in the CPT into new high efficiency heat exchangers. The installation of high efficiency heat exchangers with reasonable payback period improved FIT and reduce extra fuel gas flow rate because high efficiency heat exchangers have higher overall heat transfer coefficient and lower fouling rates than conventional shell and tube heat exchanger. This factors help to improve heat transfer efficiency between hot and cold streams in the heat exchangers. The next cost-effective option is Case 2 followed by Case 1 and Option 1. The percentage of energy and carbon dioxide emission savings as well as percentage of cost saving for Case 2 is slightly greater than Case 1. For Case 2, only parallel heat exchangers are allowed to perform online cleaning while the rest of heat exchangers severe greater fouling rates. The frequent online cleaning for parallel heat exchangers helps to improve FIT and reduce extra fuel gas flow rate. For Case 1, a few energy and carbon dioxide emission saving and percentage of cost saving is caused by little reduction of extra fuel gas compared to current practice by earlier crude preheat train overall cleaning schedule. The lowest cost-effective option is Option 1. In Option 1, a few extra fuel gas flow rate has been reduced due to some heat is recovered in furnace convection section.

Exergy analysis is conducted for the crude distillation unit. The locations and magnitudes of exergy loss are identified. From the exergy composite curve, the highest exergy loss occurs in the crude furnace as depicted by the largest gap between the exergy source and exergy sink of exergy composite curves. Exergy loss in the furnace contributes to 86% of total exergy loss in the system.

From exergy composite curve analysis, possible fuel reduction strategies are generated. From the possible fuel reduction strategies, two proposed options are implemented. The proposed options are reduction of heat loss from furnace stack and enhancement of heat recovery in the CPT. The second option is divided into three optimal cleaning schedule cases namely plant shutdown for crude preheat train overall cleaning (Case 1), heat exchangers online cleaning (Case 2) and heat exchangers online cleaning with retrofit of high efficiency heat exchangers (Case 3). Case 3 involves some investment cost due to installation of high efficiency heat exchangers. Meanwhile, Case 1 and Case 2 has no investment cost.

Comparison of energy and carbon dioxide emission savings are made for all the options. From the result, Case 3 in Option 2 is the highest energy and carbon dioxide emission saving followed by Case 2, Case 1 and Option 1. From FIT profiles, value of FIT for Case 3 at initial period is higher than FIT for Case 2. The installation of high efficiency heat exchangers at initial period for Case 3 improves FIT. Throughout the operation, FIT profile for Case 3 has smaller decay than FIT profile for Case 2. The optimal heat exchangers cleaning schedule are produced at the lowest operational cost for Case 1, Case 2 and Case 3. The optimal crude preheat train total cleaning schedule for Case 1 is on 17th and 44th months. Meanwhile for current practice, the cleaning schedule is on 28th and 49th months. Case 1 has earlier cleaning schedule than current practice. For Case 2, total number of heat exchangers online cleaning throughout 60 months period is 27. Only parallel heat exchangers are allowed to perform online cleaning namely E4 and E8. Meanwhile for Case 3, total number of heat exchangers online cleaning is 16. From the optimization result, nine heat exchangers are selected for retrofit with high efficiency heat exchangers. The selected heat exchangers are E1, E4, E5, E6, E9, E10, E11, E12, and E13. The selected heat exchangers for retrofit with high efficiency heat exchangers are able to reduce additional fuel consumption due to fouling in CPT. Thus, operational cost can be reduced and plant performance will be improved.

The economic analysis is done for all the options with and without carbon credit. From the calculation, Case 3 is the lowest percentage of operational cost saving and with and without carbon credit followed by Case 2, Case 1 and Option 1. The payback period for purchasing high efficiency heat exchangers in Case 3 is calculated. The payback period with and without carbon credit is 3 months and 4 months, respectively. The payback period is less than a year is acceptable in this study.

5.2 Future Works

This research has been carried out to develop possible fuel reduction strategies for crude distillation unit. Further study can therefore be performed by widening the scope.

The linear fouling behaviour for heat exchanger network is considered in the model formulation in heat exchangers online cleaning for Case 2 and Case 3. The typical fouling behaviours are linear or exponentially asymptotic. The further study that can be performed in this research is the implementation the exponentially asymptotic fouling model in the formulation. Lavaja & Bagajewicz (2004) and Smaili et al. (2001) had used exponentially asymptotic fouling in their heat exchanger network cleaning model.

The online heat exchangers cleaning schedules for Case 2 and Case 3 are only performing online cleaning for heat exchangers in parallel position to maintain FIT and throughput feed flow rate in the CDU. This study did not perform online cleaning for other heat exchanger in series because during the cleaning period, reduction of throughput may be necessary to maintain FIT at its desired temperature. The online cleaning for other heat exchangers is possible to implement by bypassing the heat exchangers. Lavaja & Bagajewicz (2005) considered throughput loss in their model and involve complex mathematical formulation to model throughput reduction.

The other possible future study is the simulation of heat exchanger network and planning the optimum cleaning schedule. The simulation of heat exchanger network is needed to estimate heat transfer coefficient, inlet and outlet temperatures of hot and cold streams, pressure drop as well as to obtain material properties such as specific heat and density of feed and product. In this study, the required data is taken from plant historical data in the refinery. The simulation of heat exchanger network and producing optimum cleaning schedule has been done by Sanaye & Niroomand (2007) and Ishiyama et al. (2007).

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APPENDIX A

DATA COLLECTION

A.1 Plant Data Extraction for Crude Distillation Unit

Data Extraction for CDU (Part 1)

Temperature (deg C)	Description	Design	Operation	Simulation	Agreed data
TI 966	crude inlet to E1		30.334	40	33
TI 100	crude inlet to E2	32	79.250	63	66
TIC 161	O/H Ex-C116 to E1		83.241	-	83
TI 501	kerosene outlet from E2	67	79.000	144.7	79
TI 101	crude inlet to E3	101	89.580	76.94	89
TI 402	diesel outlet from E3		117.460	140	120
TI 102	crude inlet to E4	117	95.733	101.9	100
TI 605	LSWR from E110 to E4	196.3	199.143	276	185
TI 103	crude inlet to E5	163	102.416	110.5	102
TI 300	Kerosene PA to E5		159.450	-	145
TI 165	Kerosene PA from E5 to Crude Tower		137.934	-	130
TI 104	crude inlet to E6	173.33	106.938	124	106
TI 10150	crude inlet to desalter	184.06	112.444	128	112
TI 604	LSWR from E11 to E6	242.78	224.429	276	225
TI 106	crude inlet to E7	106	120.667	128	120
TI 401	diesel from E7 to E3	175	161.300	213	170
TI 107	crude inlet to E8	143	141.778	140	160
TI 10103	Diesel PA from E8 to Crude Tower	214	210.707	-	180
TI 108	crude inlet to E9	214	183.333	193	200
TI 109	crude outlet from E9	155	191.778	200	205
TI 10117	kerosene inlet to E9		235.784	232	230
TI 110	crude from E10 to E12		225.778	228	230
TI 10121	diesel inlet to E10		291.662	285	279
TI 500	kerosene from E9 to E2	239	192.222	201	205
TI 400	diesel from E10 to E7	288	226.350	247	235
TI 111	crude outlet from E11	184.77	197.000	211	215
TI 603	LSWR from E11 to E6	209.17	230.000	276	245
TIC 160	crude inlet to Pre-flash column		201.555	201	203

Data Extraction for CDU (Part 2)

Temperature (deg C)	Description	Design	Operation	Simulation	Agreed data
TI 602	LSWR from E12 to E11	333	242.200	-	242
TI 168	bottom pre-flash to E13		198.907	199.1	200
TI 10120	LSWR to E13		339.788	347	347
TI 600	LSWR from E13 to E12		299.214	302	300
TI 179	furnace inlet temperature FIT		209.579	220	215
TC 104	tower inlet temperature (TIT)		362.344	367.1	367
TI 10114 A/B	furnace Stack temperature		292.626	-	290
Flowrate (m3/hr)	Description	Design	Operation	Simulation	Agreed data
FIC 101	Crude from storage to CDU				
	Kbpsd	40	49.300	46.1	50
	m3/hr	264.98	326.742	305.397	331.224
FI 104	LSWR flowrate from E4 to storage	-	61.191	-	70
FI 110	offgas flowrate to furnace	-	3960.377	-	4161
FC 111	Kerosene PA to E5	-	90.933	-	90
FI 162	crude flowrate to E10	-	185.366	-	185
FI 173	bottom flowrate of pre-flash column to E13	-	279.02	-	300
FC 102	Kerosene to storage	-	74.30	-	71
FC 103	diesel exit to E10	-	60.00	-	104.44
FC 112	Diesel Pump around to E8	-	223.20	-	170.15
offgas (mol%)	Description	Design	Operation	Simulation	Agreed data
OffgasC402.Hydrogen	Hydrogen	-	52.052	N/A	52.6
OffgasC402.Ethane	Ethane	-	9.379	-	8.75
OffgasC402.Propane	Propane	-	5.450	-	5.63
OffgasC402.i-Butane	i-Butane	-	1.064	-	0.94
OffgasC402.i-Pentane	i-Pentane	-	0.194	-	0.16
OffgasC402.Methane	Methane	-	30.145	-	29.76
OffgasC402.n-Butane	n-Butane	-	0.841	-	1.01
OffgasC402.n-Pentane	n-Pentane	-	0.095	-	0.16
OffgasC402.Pent/Hex&Heavier	Pent/Hex & Heavier	-	0.779	-	0.98

Data Extraction for CDU (Part 3)

Cp cold stream (kJ/kg C)	Values	Agreed data
E1	2.14	2.0024
E2	2.185	2.0024
E3	2.4084	2.0024
E4	2.4084	2.0024
E5	2.55	2.0024
E6	2.566	2.0024
E7	2.585	2.0024
E8	2.7307	2.0024
E9	2.7406	2.0024
E10	2.8471	2.0024
E11	2.973	2.0024
E12	2.8471	2.0024
E13	2.7902	2.0024
furnace	2.83	2.0024
Density (kg/L)	Values	Agreed data
Kerosene	0.7864	0.791
Tapis crude	0.8032	0.815
Diesel	0.8312	0.828
LSWR	0.8722	0.87
Overhead vapor from pre-flash column	0.6473	0.6473

Data of heat exchanger network for Case 2 and Case 3

	Heat exchanger												
Parameters	E1	E2	E3	E4	E5	E6	E7	E8	E9	E10	E11	E12	E13
Tc1 (^{o}C)	33	66	89	100	102	106	120	160	200	205	205	223	200
Tc2 (^{o}C)	66	89	100	102	106	112	160	200	205	230	215	232	215
Th1 (^{o}C)	83	205	170	185	145	225	235	205	230	279	242	300	347
Th2 (^{o}C)	40	79	120	179	130	185	170	180	205	235	225	242	300
Fh (kg/s)	16	15.6	24	17	19.8	17	24	39.1	15.6	24	17	17	17
Fc (kg/s)	75	75	75	75	75	75	75	75	75	42	33	75	68
Cc (kJ/kg^{o}C)	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Ch (kJ/kg^{o}C)	7.2	1.8	1.4	3.0	2.0	1.3	3.9	6.2	1.9	2.0	3.4	1.4	3.4

A.2 Data of fouling resistance for crude preheat train

t	Rf for crude preheat train (m^{2o}C/kW)												
	E1	E2	E3	E4	E5	E6	E7	E8	E9	E10	E11	E12	E13
1	0.030	0.048	0.013	0.012	0.011	0.033	0.062	0.028	0.018	0.023	0.015	0.014	0.046
2	0.065	0.112	0.067	0.015	0.015	0.066	0.133	0.069	0.029	0.041	0.027	0.029	0.087
3	0.106	0.174	0.079	0.030	0.034	0.098	0.203	0.088	0.041	0.063	0.041	0.045	0.125
4	0.122	0.237	0.112	0.051	0.047	0.127	0.278	0.135	0.054	0.089	0.058	0.060	0.165
5	0.169	0.238	0.145	0.070	0.058	0.159	0.379	0.178	0.068	0.106	0.068	0.083	0.207
6	0.211	0.287	0.173	0.079	0.073	0.189	0.413	0.210	0.076	0.127	0.076	0.094	0.257
7	0.260	0.427	0.203	0.092	0.083	0.232	0.480	0.190	0.087	0.141	0.084	0.110	0.306
8	0.290	0.479	0.236	0.104	0.095	0.247	0.611	0.227	0.095	0.164	0.097	0.128	0.317
9	0.299	0.484	0.277	0.118	0.112	0.261	0.699	0.261	0.101	0.184	0.113	0.133	0.371
10	0.336	0.596	0.298	0.122	0.112	0.289	0.776	0.289	0.114	0.203	0.127	0.147	0.399

APPENDIX B

CALCULATION OF SOME PARAMETERS

B.1 Theoretical Flame Temperature (TFT) calculation

The off gas in furnace's fuel gas contains 52.6 mol% hydrogen, 29.76 mol% methane, 8.75 mol% ethane, 5.63 mol% propane, 1.01 mol% n-butane, 0.98 mol% hexane, 0.94 mol% i-butane, 0.16 mol% i-pentane and 0.16 mol% n-pentane. The combustion is at 10% excess air and the ambient temperature, T_o is 27°C.

The standard heat of combustion for all components in off gas is:

$CH_4+2O_2 \rightarrow CO_2+2H_2O$	$\Delta H^o c$= -802,600 kJ/kmol
$C_2H_6+7/2O_2 \rightarrow 2CO_2+3H_2O$	$\Delta H^o c$= -1,428,600 kJ/kmol
$C_3H_8+5O_2 \rightarrow 3CO_2+4\ H_2O$	$\Delta H^o c$= -2,043,100 kJ/kmol
$C_4H_{10}+13/2O_2 \rightarrow 4CO_2+5\ H_2O$	$\Delta H^o c$= -2,657,300 kJ/kmol
$C_5H_{12}+8O_2 \rightarrow 5CO_2+6\ H_2O$	$\Delta H^o c$= -3,244,900 kJ/kmol
$C_6H_{14}+19O_2 \rightarrow 6CO_2+7\ H_2O$	$\Delta H^o c$= -3,855,100 kJ/kmol
$H_2+ O_2 \rightarrow H_2O$	$\Delta H^o c$= -241,800 kJ/kmol

Amount of Reactants in kmol are,

H_2 = 0.526

CH_4 = 0.298

C_2H_6 = 0.0875

C_3H_8 = 0.0563

C_4H_{10} = 0.0195

C_5H_{12} = 0.0032

C_6H_{14} = 0.0098

O_2 = [0.526+2(0.298)+7/2(0.0875)+5(0.056)+13/2(0.0195)+8(0.0032)+19(0.0098)][1.1]

= 2.251

N_2 =2.251 × (79/21) = 8.468

Amount of Products in kmol are,

CO_2 = 0.298+2(0.0875)+3(0.056)+4(0.0195)+5(0.0032)+6(0.0098)

=0.7938

H_2O =0.526+2(0.298)+3(0.0875)+4(0.056)+5(0.0195)+6(0.0032)+7(0.0098)

= 1.7938

N_2 =8.468

Excess O_2= 2.251-0.526-2(0.298)-7/2(0.0875)-5(0.056)-13/2(0.0195)-8(0.0032)-19(0.0098)

= 0.2042

Assume TFT is 1800 Kelvin. Specific heat, Cp for CO_2, H_2O, N_2 and O_2 are calculated.

Cp for CO_2

$= 0.2937\times 10^5+0.3454\times 10^5[(1.428\times 10^3/T)/\sinh(1.428\times 10^3/T)]^2+0.264\times 10^5[(588/T)/\cosh(588/T)]^2$

= 60.029 kJ/kmol K

Cp for H_2O

$= 0.3336\times 10^5+0.2679\times 10^5[(2.6106\times 10^3/T)/\sinh(2.6106\times 10^3/T)]^2+0.089\times 10^5[(1169/T)/\cosh(1169/T)]^2$

= 49.768 kJ/kmol K

Cp for N_2

$= 0.2911\times 10^5+0.0861\times 10^5[(1.7016\times 10^3/T)/\sinh(1.7016\times 10^3/T)]^2+0.001\times 10^5[(909.8/T)/\cosh(909.8/T)]^2$

= 35.575 kJ/kmol K

Cp for O_2

$= 0.2910\times 10^5+0.1004\times 10^5[(2.527\times 10^3/T)/\sinh(2.527\times 10^3/T)]^2+0.094\times 10^5[(1153.8/T)/\cosh(1153.8/T)]^2$

= 34.537 kJ/kmol K

TFT is calculated using energy balance.

$$\int_{T_{inlet}}^{T_o} \Delta H_{reactant} + \Delta H_C^0 + \int_{T_0}^{TFT} \Delta H_{product} = 0$$

The combustion system is adiabatic where T_o is equal to T_{inlet} for $\Delta H_{reactant}$. Thus, $\Delta H_{reactant}$ is zero. The equation becomes

$$\Delta H_C^0 + \int_{T_0}^{TFT} \Delta H_{product} = 0$$

$\Delta H^o c$ = (-802600 kJ/kmol)(0.298 kmol)+(-241800 kJ/kmol)(0.526 kmol)+(-1428600 kJ/kmol)(0.0875 kmol)+(-2043100 kJ/kmol)(0.056 kmol)+(-2657300 kJ/kmol)(0.0195 kmol)+(-3244900 kJ/kmol)(0.0032 kmol)+(-3855100 kJ/kmol)(0.0098 kmol)

= -705758.71 kJ

$\Delta H_{product}$ = [(0.7938 kmol)(60.029 kJ/kmol K)+(1.7938 kmol)(49.768 kJ/kmol K)+(8.468 kmol)(35.575 kJ/kmol K)+(0.2042 kJ/kmol K)(34.537 kmol)](TFT-300)

= 445.23 kJ/K (TFT-300)

Substitute in $\Delta H^o c$ and $\Delta H_{product}$ into energy balance,

0 = -705758.71 kJ + 445.23 kJ/K (TFT-300)

TFT = 1888 Kelvin ≈ 1615°C

The percentage error is 4 % (The value of TFT is acceptable)

B.2 Specific heat (Cp) calculation of fuel gas in CDU furnace

The equation to calculate Cp for every component in fuel gas is as below. Fuel gas consists of hydrogen, methane, ethane, propane, n-butane, hexane, i-butane, i-pentane and n-pentane. The ambient temperature is 300 Kelvin.

$$Cp = C1 + C2\left[\frac{C3/T}{\sinh\left(\frac{C3}{T}\right)}\right]^2 + C4\left[\frac{C5/T}{\cosh\left(\frac{C5}{T}\right)}\right]^2$$

Components	C1	C2	C3	C4	C5	Cp(J/kmol K)
hydrogen	27620	9560	2466	3760	567	28792.21
ethane	40330	134220	1655	73220	753	52621.06
propane	51920	192450	1626.5	116800	724	73821.08
I butane	65490	247760	1587	157500	-707	97041.19
I pentane	74600	326500	1545	192300	-666.7	119338.8
methane	3300000	79930	2086.9	41600	992	3302450
n butane	71340	243000	1630	150330	730.42	98839.12
n pentane	88050	301100	1650.2	189200	748	120349.8
hexane	104400	352300	1694.6	236900	762	142514.8

Cp is converted from J/kmol K to kJ/kg K by using the following equations.

$$Cp\left(\frac{J}{kmolK}\right)\times kmol\%\times\frac{1}{MW\left(\frac{kg}{kmol}\right)}\times\frac{1kJ}{1000J}=Cp\left(\frac{kJ}{kgK}\right)$$

Components	MW (kg/kmol)	kmol%	Cp (kJ/kg K)
hydrogen	2	0.526	7.572
ethane	30	0.0875	0.153
propane	44	0.056	0.094
I butane	58.123	0.0094	0.016
I pentane	72.15	0.0016	0.003
methane	16	0.2976	61.426
n butane	58.123	0.0101	0.017
n pentane	72.15	0.0016	0.003
hexane	86.177	0.0098	0.016
Cp Total (kJ/kg K)			69.3

APPENDIX C

LIST OF SELECTED VALUES FROM OPTIMAL SOLUTION IN OPTION 2

C.1 Extra fuel gas flow rate for current practice and all cases

Months	Current Practice	Case 1	Case 2	Case 3
0	0.00	0.00	0.00	0.00
1	37.21	17.93	28.03	20.37
2	73.00	35.86	11.89	41.17
3	107.37	53.78	40.72	62.20
4	140.33	71.71	70.15	8.49
5	171.87	89.64	99.90	29.82
6	202.00	107.57	34.42	51.49
7	230.71	125.50	64.74	73.27
8	258.00	143.42	45.15	95.02
9	283.87	161.35	76.16	17.93
10	308.33	179.28	107.59	40.31
11	331.37	197.21	60.65	62.91
12	352.99	215.14	92.63	85.52
13	373.20	233.06	124.96	24.78
14	391.99	250.99	75.51	47.89
15	409.37	268.92	108.43	71.15
16	425.32	286.85	141.61	94.36
17	439.86	0.00	89.82	31.15
18	452.99	17.93	123.61	54.91
19	464.70	35.86	157.60	78.77
20	474.99	53.78	103.62	102.51
21	483.86	71.71	138.23	37.11
22	491.32	89.64	112.56	61.47
23	497.36	107.57	147.70	85.87
24	501.98	125.50	121.31	110.10
25	505.19	143.42	156.96	42.76
26	506.98	161.35	192.67	67.65
27	507.35	179.28	134.08	92.55

Months	Current Practice	Case 1	Case 2	Case 3
28	0.00	197.21	170.47	46.80
29	37.21	215.14	142.39	72.07
30	73.00	233.06	179.25	97.31
31	107.37	250.99	216.12	50.71
32	140.33	268.92	154.56	76.33
33	171.87	286.85	192.11	101.91
34	202.00	304.78	162.48	54.50
35	230.71	322.70	200.48	80.46
36	258.00	340.63	170.27	106.34
37	283.87	358.56	208.68	131.92
38	308.33	376.49	177.90	59.38
39	331.37	394.41	216.75	85.76
40	352.99	412.34	255.48	61.76
41	373.20	430.27	189.13	88.33
42	391.99	448.20	228.59	114.78
43	409.37	466.13	196.46	65.25
44	425.32	0.00	236.33	92.10
45	439.86	17.93	203.68	118.82
46	452.99	35.86	243.93	68.66
47	464.70	53.78	210.78	95.78
48	474.99	71.71	251.42	122.76
49	0.00	89.64	291.81	149.36
50	37.21	107.57	221.21	73.08
51	73.00	125.50	262.40	100.56
52	107.37	143.42	303.35	127.86
53	140.33	161.35	231.43	76.33
54	171.87	179.28	273.16	104.05
55	202.00	197.21	314.62	131.57
56	230.71	215.14	241.43	79.51
57	258.00	233.06	244.72	107.46
58	283.87	250.99	287.15	135.21
59	308.33	268.92	329.27	162.53
60	0.00	0.00	0.00	0.00

C.2 FIT for current practice and all cases

Months	Current Practice	Case 1	Case 2	Case 3
0	215.00	215.00	215.09	226.74
1	214.44	214.44	214.21	226.11
2	213.88	213.88	214.72	225.46
3	213.32	213.32	213.82	224.80
4	212.76	212.76	212.90	226.48
5	212.20	212.20	211.97	225.81
6	211.64	211.64	214.01	225.13
7	211.08	211.08	213.07	224.45
8	210.52	210.52	213.68	223.77
9	209.96	209.96	212.71	226.18
10	209.40	209.40	211.73	225.48
11	208.84	208.84	213.19	224.78
12	208.28	208.28	212.19	224.07
13	207.72	207.72	211.18	225.97
14	207.16	207.16	212.73	225.25
15	206.60	206.60	211.70	224.52
16	206.04	206.04	210.66	223.79
17	205.47	215.00	212.28	225.77
18	204.91	214.44	211.23	225.03
19	204.35	213.88	210.16	224.28
20	203.79	213.32	211.85	223.54
21	203.23	212.76	210.77	225.58
22	202.67	212.20	211.57	224.82
23	202.11	211.64	210.47	224.06
24	201.55	211.08	211.30	223.30
25	200.99	210.52	210.18	225.41
26	200.43	209.96	209.07	224.63
27	199.87	209.40	210.90	223.85
28	215.00	208.84	209.76	225.28
29	214.44	208.28	210.64	224.49
30	213.88	207.72	209.49	223.70
31	213.32	207.16	208.34	225.16
32	212.76	206.60	210.26	224.36
33	212.20	206.04	209.09	223.56
34	211.64	205.48	210.01	225.04

Months	Current Practice	Case 1	Case 2	Case 3
35	211.08	204.92	208.82	224.23
36	210.52	204.35	209.77	223.42
37	209.96	203.79	208.57	222.62
38	209.40	203.23	209.53	224.89
39	208.84	202.67	208.32	224.06
40	208.28	202.11	207.11	224.81
41	207.72	201.55	209.18	223.98
42	207.16	200.99	207.95	223.15
43	206.60	200.43	208.95	224.70
44	206.04	215.00	207.70	223.86
45	205.47	214.44	208.72	223.03
46	204.91	213.88	207.47	224.60
47	204.35	213.32	208.50	223.75
48	203.79	212.76	207.23	222.91
49	215.00	212.20	205.97	222.07
50	214.44	211.64	208.18	224.46
51	213.88	211.08	206.89	223.60
52	213.32	210.52	205.61	222.75
53	212.76	209.96	207.86	224.36
54	212.20	209.40	206.55	223.49
55	211.64	208.84	205.26	222.63
56	211.08	208.28	207.54	224.26
57	210.52	207.72	207.44	223.38
58	209.96	207.16	206.12	222.52
59	209.40	206.60	204.80	221.66
60	215.00	215.00	215.00	215.00

APPENDIX D

MODEL FORMULATION FOR OPTION 2 IN GAMS SOFTWARE

D.1 Case 1: crude preheat train total cleaning

```
SET
t months /0*59/;

POSITIVE VARIABLE

FIT(t)    Furnace inlet temperature
FG(t)     fuel gas consumption
CO2(t)
;

VARIABLE

z         objective function minimize cost
;

BINARY VARIABLE

y(t)      heat exchanger is cleaned at period t
;

EQUATIONS

cost              minimize cost
FuelFIT           FG vs FIT correlation
initialFIT        initial fuel gas
Fuel              fuel gas correlation
lowerFIT          lower FIT
CO2emission       carbon dioxide emission
;

Fuel(t)..         FIT(t)    =e= (FIT(t-1)-0.5603)*(1-y(t))+(FIT('0')*y(t));
FuelFIT(t)..      FG(t)     =e=  -31.997*FIT(t)+10139;
initialFIT('0')..  FIT('0')  =e= 215;
lowerFIT(t)..     FIT(t)    =g= 200;
CO2emission(t)..  CO2(t)    =e= 1.32*FG(t);

cost..            z         =e= sum(t,(FG(t)-FG('0')))*0.029*14.549*720 +
sum(t,y(t))*(520000+5450000);


Model scheduling /all/;
Solve scheduling using MINLP minimize z;
Display z.l,y.l
```

D.2 Case 2: heat exchangers online cleaning

```
SET
i heat exchangers /E1,E2,E3,E4,E5,E6,E7,E8,E9,E10,E11,E12,E13/
t months /0*59/;

PARAMETERS

Fh(i)    hot stream flowrate

        /       E1 = 16
                E2 = 15.6
                E3 = 24
                E4 = 17
                E5 = 19.8
                E6 = 17
                E7 = 24
                E8 = 39.1
                E9 = 15.6
                E10 = 24
                E11 = 17
                E12 = 17
                E13 = 17       /

Fc(i)    cold stream flowrate

        /       E1 = 75
                E2 = 75
                E3 = 75
                E4 = 75
                E5 = 75
                E6 = 75
                E7 = 75
                E8 = 75
                E9 = 75
                E10 = 42
                E11 = 33
                E12 = 75
                E13 = 68  /

Ch(i)    specific heat of hot stream

        /       E1 = 7.21
                E2 = 1.76
                E3 = 1.38
                E4 = 2.95
                E5 = 2.02
                E6 = 1.33
                E7 = 3.86
                E8 = 6.15
                E9 = 1.93
                E10 = 1.99
                E11 = 3.37
                E12 = 1.37
                E13 = 3.41     /

Cc(i)    specific heat of cold stream
```

/ E1 = 2.0024
E2 = 2.0024
E3 = 2.0024
E4 = 2.0024
E5 = 2.0024
E6 = 2.0024
E7 = 2.0024
E8 = 2.0024
E9 = 2.0024
E10 = 2.0024
E11 = 2.0024
E12 = 2.0024
E13 = 2.0024 /

Uc(i) heat transfer coefficient at clean condition

/ E1 = 1.073
E2 = 0.66
E3 = 0.095
E4 = 0.033
E5 = 0.1
E6 = 0.085
E7 = 0.736
E8 = 4.47
E9 = 0.542
E10 = 0.32
E11 = 0.185
E12 = 0.212
E13 = 0.227 /

A(i) heat transfer area of heat exchangers

/ E1 = 411.5
E2 = 111
E3 = 361.8
E4 = 111
E5 = 180.91
E6 = 111.39
E7 = 132.5
E8 = 125
E9 = 111.39
E10 = 169.91
E11 = 153.66
E12 = 165.5
E13 = 78.2 /

R(i) heat transfer ratio

/ E1 = 1.303
E2 = 5.478
E3 = 4.545
E4 = 3
E5 = 3.75
E6 = 6.67
E7 = 1.63
E8 = 0.625
E9 = 5
E10 = 1.76

```
        E11 = 1.7
        E12 = 6.44
        E13 = 3.133    /

POSITIVE VARIABLE

Tc1(i,t)        inlet cold stream temperature
Th2(i,t)        outlet hot stream temperature
Tc2(i,t)        outlet cold stream temperature
FG(t)           fuel gas consumption
Rf(i,t)         total fouling factor
dRf(i,t)        fouling rates
Uf(i,t)         fouled heat transfer coefficient
Th1(i,t)        inlet hot stream temperature
CO2emission     carbon dioxide emission
;

VARIABLE
z          objective function minimize cost
;

BINARY VARIABLE

y(i,t)       ith heat exchanger is cleaned at period t;

EQUATIONS

cost            minimize cost
foulingUf       Rf and U correlation
outletTh        correlation of hot outlet temperature
outletTc        correlation of cold outlet temperature
fouling         fouling rate
Fuel            correlation of fuel gas
iRf             initial fouling for all heat exchanger at time zero
CO2emission     carbon dioxide emission
dRfE1           fouling rate E1
dRfE2           fouling rate E2
dRfE3           fouling rate E3
dRfE4           fouling rate E4
dRfE5           fouling rate E5
dRfE6           fouling rate E6
dRfE7           fouling rate E7
dRfE8           fouling rate E8
dRfE9           fouling rate E9
dRfE10          fouling rate E10
dRfE11          fouling rate E11
dRfE12          fouling rate E12
dRfE13          fouling rate E13
inletTcE1       inlet cold stream temperature for E1
inletTcE2       inlet cold stream temperature for E2
inletTcE3       inlet cold stream temperature for E3
inletTcE4       inlet cold stream temperature for E4
inletTcE5       inlet cold stream temperature for E5
inletTcE6       inlet cold stream temperature for E6
inletTcE7       inlet cold stream temperature for E7
inletTcE8       inlet cold stream temperature for E8
inletTcE9       inlet cold stream temperature for E9
inletTcE10      inlet cold stream temperature for E10
```

```
inletTcE11          inlet cold stream temperature for E11
inletTcE12          inlet cold stream temperature for E12
inletTcE13          inlet cold stream temperature for E13
inletThE1           inlet hot stream temperature for E1
inletThE2           inlet hot stream temperature for E2
inletThE3           inlet hot stream temperature for E3
inletThE4           inlet hot stream temperature for E4
inletThE5           inlet hot stream temperature for E5
inletThE6           inlet hot stream temperature for E6
inletThE7           inlet hot stream temperature for E7
inletThE8           inlet hot stream temperature for E8
inletThE9           inlet hot stream temperature for E9
inletThE10          inlet hot stream temperature for E10
inletThE11          inlet hot stream temperature for E11
inletThE12          inlet hot stream temperature for E12
inletThE13          inlet hot stream temperature for E13
consE1              binary variable y constraint for E1
consE2              binary variable y constraint for E2
consE3              binary variable y constraint for E3
consE4              binary variable y constraint for E4
consE5              binary variable y constraint for E5
consE6              binary variable y constraint for E6
consE7              binary variable y constraint for E7
consE8              binary variable y constraint for E8
consE9              binary variable y constraint for E9
consE10             binary variable y constraint for E10
consE11             binary variable y constraint for E11
consE12             binary variable y constraint for E12
consE13             binary variable y constraint for E13
cons                FIT constraint
consRfE4            fouling constraint for E4
consRfE8            fouling constraint for E8
;

inletTcE1('E1',t)..          Tc1('E1',t)    =e= 33;
inletTcE2('E2',t)..          Tc1('E2',t)    =e= Tc2('E1',t);
inletTcE3('E3',t)..          Tc1('E3',t)    =e= Tc2('E2',t);
inletTcE4('E4',t)..          Tc1('E4',t)    =e= Tc2('E3',t);
inletTcE5('E5',t)..          Tc1('E5',t)    =e= Tc2('E4',t);
inletTcE6('E6',t)..          Tc1('E6',t)    =e= Tc2('E5',t);
inletTcE7('E7',t)..          Tc1('E7',t)    =e= Tc2('E6',t);
inletTcE8('E8',t)..          Tc1('E8',t)    =e= Tc2('E7',t);
inletTcE9('E9',t)..          Tc1('E9',t)    =e= Tc2('E8',t);
inletTcE10('E10',t)..        Tc1('E10',t)   =e= Tc2('E9',t);
inletTcE11('E11',t)..        Tc1('E11',t)   =e= Tc2('E9',t);
inletTcE12('E12',t)..        Tc1('E12',t)   =e= 0.56*Tc2('E10',t)+0.44*Tc2('E11',t);
inletTcE13('E13',t)..        Tc1('E13',t)   =e= Tc2('E12',t)-32;


inletThE1('E1',t)..          Th1('E1',t)    =e= 83;
inletThE2('E2',t)..          Th1('E2',t)    =e= Th2('E9',t);
inletThE3('E3',t)..          Th1('E3',t)    =e= Th2('E7',t);
inletThE4('E4',t)..          Th1('E4',t)    =e= Th2('E6',t);
inletThE5('E5',t)..          Th1('E5',t)    =e= 145;
inletThE6('E6',t)..          Th1('E6',t)    =e= Th2('E11',t);
inletThE7('E7',t)..          Th1('E7',t)    =e= Th2('E10',t);
inletThE8('E8',t)..          Th1('E8',t)    =e= 205;
inletThE9('E9',t)..          Th1('E9',t)    =e= 230;
```

```
inletThE10('E10',t)..         Th1('E10',t)  =e= 279;
inletThE11('E11',t)..         Th1('E11',t)  =e= Th2('E12',t);
inletThE12('E12',t)..         Th1('E12',t)  =e= Th2('E13',t);
inletThE13('E13',t)..         Th1('E13',t)  =e= 347;


Fuel(t)..            FG(t)               =e=  -31.997*Tc2(‘E13’,t)+10139;
iRf(i,'0')..         Rf(i,'0')           =e= 0;
fouling(i,t)..       Rf(i,t)             =e= (Rf(i,t-1)+dRf(i,t))*(1-y(i,t))+(Rf(i,'0')*y(i,t));
foulingUf(i,t)..     Uf(i,t)             =e= 1/((1/Uc(i))+Rf(i,t));
outletTh(i,t)..      Th2(i,t)            =e= ((R(i)-1)*Th1(i,t)+(exp(((Uf(i,t)*A(i))/(Fc(i)*Cc(i)))*(R(i)-
1))-1)*R(i)*Tc1(i,t))/(R(i)*exp(((Uf(i,t)*A(i))/(Fc(i)*Cc(i)))*(R(i)-1))-1);
outletTc(i,t)..      Tc2(i,t)            =e= Tc1(i,t)+(Th1(i,t)-Th2(i,t))/R(i);
cons('E13',t)..      Tc2('E13',t)        =g= 200;
CO2emission(t)   CO2(t)              =e= 1.32*FG(t);

dRfE1('E1',t)..     dRf('E1',t)          =e= 0.035;
dRfE2('E2',t)..     dRf('E2',t)          =e= 0.06;
dRfE3('E3',t)..     dRf('E3',t)          =e= 0.033;
dRfE4('E4',t)..      dRf('E4',t)         =e= 0.013;
dRfE5('E5',t)..     dRf('E5',t)          =e= 0.012;
dRfE6('E6',t)..     dRf('E6',t)          =e= 0.032;
dRfE7('E7',t)..     dRf('E7',t)          =e= 0.07;
dRfE8('E8',t)..     dRf('E8',t)          =e= 0.03;
dRfE9('E9',t)..     dRf('E9',t)          =e= 0.011;
dRfE10('E10',t).. dRf('E10',t)         =e= 0.02;
dRfE11('E11',t).. dRf('E11',t)         =e= 0.012;
dRfE12('E12',t).. dRf('E12',t)         =e= 0.015;
dRfE13('E13',t).. dRf('E13',t)         =e= 0.04;

consE1('E1')..      sum(t,y('E1',t))     =l= 1;
consE2('E2')..      sum(t,y('E2',t))     =l= 1;
consE3('E3')..      sum(t,y('E3',t))     =l= 1;
consE4('E4')..      sum(t,y('E4',t))     =l= 60;
consE5('E5')..      sum(t,y('E5',t))     =l= 1;
consE6('E6')..      sum(t,y('E6',t))     =l= 1;
consE7('E7')..      sum(t,y('E7',t))     =l= 1;
consE8('E8')..      sum(t,y('E8',t))     =l= 60;
consE9('E9')..      sum(t,y('E9',t))     =l= 1;
consE10('E10')..  sum(t,y('E10',t))  =l= 1;
consE11('E11')..  sum(t,y('E11',t))  =l= 1;
consE12('E12')..  sum(t,y('E12',t))  =l= 1;
consE13('E13')..  sum(t,y('E13',t))  =l= 1;

consRfE4('E4',t)..   Rf('E4',t) =l= 0.221;
consRfE8('E8',t)..   Rf('E8',t) =l= 0.51;

cost..                          z                =e= sum(t,(FG(t)-FG('0')))*0.029*14.549*720 +
sum((i,t),40000*y(i,t));

Model scheduling /all/;
Solve scheduling using MINLP minimize z;
Display z.l,y.l;
```

D.3 Case 3: heat exchangers online cleaning with retrofit of high efficiency heat exchangers

```
SET
i    heat exchangers /E1,E2,E3,E4,E5,E6,E7,E8,E9,E10,E11,E12,E13/
t    time period     /0*59/
m(t)  /0*59/
;

PARAMETERS

Fh(i)    hot stream flowrate

        /       E1 = 16
                E2 = 15.6
                E3 = 24
                E4 = 17
                E5 = 19.8
                E6 = 17
                E7 = 24
                E8 = 39.1
                E9 = 15.6
                E10 = 24
                E11 = 17
                E12 = 17
                E13 = 17       /

Fc(i)    cold stream flowrate

        /       E1 = 75
                E2 = 75
                E3 = 75
                E4 = 75
                E5 = 75
                E6 = 75
                E7 = 75
                E8 = 75
                E9 = 75
                E10 = 42
                E11 = 33
                E12 = 75
                E13 = 68  /

Ch(i)    specific heat of hot stream

        /       E1 = 7.21
                E2 = 1.76
                E3 = 1.38
                E4 = 2.95
                E5 = 2.02
                E6 = 1.33
                E7 = 3.86
                E8 = 6.15
                E9 = 1.93
                E10 = 1.99
                E11 = 3.37
                E12 = 1.37
                E13 = 3.41     /
```

Cc(i) specific heat of cold stream

/ E1 = 2.0024
E2 = 2.0024
E3 = 2.0024
E4 = 2.0024
E5 = 2.0024
E6 = 2.0024
E7 = 2.0024
E8 = 2.0024
E9 = 2.0024
E10 = 2.0024
E11 = 2.0024
E12 = 2.0024
E13 = 2.0024 /

Uc(i) heat transfer coefficient at clean condition

/ E1 = 1.073
E2 = 0.66
E3 = 0.095
E4 = 0.033
E5 = 0.1
E6 = 0.085
E7 = 0.736
E8 = 4.47
E9 = 0.542
E10 = 0.32
E11 = 0.185
E12 = 0.212
E13 = 0.227 /

A(i) heat transfer area of heat exchangers

/ E1 = 411.5
E2 = 111
E3 = 361.8
E4 = 111
E5 = 180.91
E6 = 111.39
E7 = 132.5
E8 = 125
E9 = 111.39
E10 = 169.91
E11 = 153.66
E12 = 165.5
E13 = 78.2 /

R(i) heat transfer ratio

/ E1 = 1.303
E2 = 5.478
E3 = 4.545
E4 = 3
E5 = 3.75
E6 = 6.67
E7 = 1.63
E8 = 0.625

```
                E9 = 5
                E10 = 1.76
                E11 = 1.7
                E12 = 6.44
                E13 = 3.133    /

dRfhex(i)       difference in fouling for high efficiency heat exchanger (90%)


/               E1 = 0.0315
                E2 = 0.054
                E3 = 0.0297
                E4 = 0.0117
                E5 = 0.0108
                E6 = 0.0288
                E7 = 0.063
                E8 = 0.027
                E9 = 0.0099
                E10 = 0.018
                E11 = 0.0108
                E12 = 0.0135
                E13 = 0.036    /


POSITIVE VARIABLE

Rf(i,t)             fouling factor of ith heat exchanger at period t
dRf(i,t)            fouling rate for ith heat exchanger at period t
dRfhhx(i,t)         difference in fouling rate for high efficiency heat exchanger
Tc1(i,t)            inlet cold temperature
Tc2(i,t)            outlet cold temperature
Th1(i,t)            inlet hot stream temperature
Th2(i,t)            outlet hot temperature
FG(t)               fuel gas consumption
Uf(i,t)             fouled heat transfer coefficient
CO2(t)
;

VARIABLE
z           objective function minimize cost
;

BINARY VARIABLE

y(i,t)      ith heat exchanger is cleaned in period t
h(i,t)      ith heat exchanger is changed in period t
;

EQUATIONS

cost                minimize cost
foulingUf           Rf and U correlation
outletTh            corelation of Th outlet
outletTc            correlation of Tc outlet
foulingy            fouling rate for y binary variable
FuelFIT             FG vs FIT
iRf                 initial fouling for all heat exchanger
CO2emission         carbon dioxide emission
```

```
dRfE1           fouling rate E1
dRfE2           fouling rate E2
dRfE3           fouling rate E3
dRfE4           fouling rate E4
dRfE5           fouling rate E5
dRfE6           fouling rate E6
dRfE7           fouling rate E7
dRfE8           fouling rate E8
dRfE9           fouling rate E9
dRfE10          fouling rate E10
dRfE11          fouling rate E11
dRfE12          fouling rate E12
dRfE13          fouling rate E13
inletTcE1       inlet cold stream temperature for E1
inletTcE2       inlet cold stream temperature for E2
inletTcE3       inlet cold stream temperature for E3
inletTcE4       inlet cold stream temperature for E4
inletTcE5       inlet cold stream temperature for E5
inletTcE6       inlet cold stream temperature for E6
inletTcE7       inlet cold stream temperature for E7
inletTcE8       inlet cold stream temperature for E8
inletTcE9       inlet cold stream temperature for E9
inletTcE10      inlet cold stream temperature for E10
inletTcE11      inlet cold stream temperature for E11
inletTcE12      inlet cold stream temperature for E12
inletTcE13      inlet cold stream temperature for E13
inletThE1       inlet hot stream temperature for E1
inletThE2       inlet hot stream temperature for E2
inletThE3       inlet hot stream temperature for E3
inletThE4       inlet hot stream temperature for E4
inletThE5       inlet hot stream temperature for E5
inletThE6       inlet hot stream temperature for E6
inletThE7       inlet hot stream temperature for E7
inletThE8       inlet hot stream temperature for E8
inletThE9       inlet hot stream temperature for E9
inletThE10      inlet hot stream temperature for E10
inletThE11      inlet hot stream temperature for E11
inletThE12      inlet hot stream temperature for E12
inletThE13      inlet hot stream temperature for E13
consE1          binary variable y constraint for E1
consE2          binary variable h and y constraint for E2
consE3          binary variable h and y constraint for E3
consE4          binary variable h and y constraint for E4
consE5          binary variable h and y constraint for E5
consE6          binary variable h and y constraint for E6
consE7          binary variable h and y constraint for E7
consE8          binary variable h and y constraint for E8
consE9          binary variable h and y constraint for E9
consE10         binary variable h and y constraint for E10
consE11         binary variable h and y constraint for E11
consE12         binary variable h and y constraint for E12
consE13         binary variable h and y constraint for E13
foulinghex
cons1           constraint for FIT
cons2           constraint for two binary variables
cons3           constraint for selecting variable y at all period t
cons4           constraint for selecting variable y at period t=0
;
```

```
inletTcE1('E1',t)..          Tc1('E1',t)    =e= 33;
inletTcE2('E2',t)..          Tc1('E2',t)    =e= Tc2('E1',t);
inletTcE3('E3',t)..          Tc1('E3',t)    =e= Tc2('E2',t);
inletTcE4('E4',t)..          Tc1('E4',t)    =e= Tc2('E3',t);
inletTcE5('E5',t)..          Tc1('E5',t)    =e= Tc2('E4',t);
inletTcE6('E6',t)..          Tc1('E6',t)    =e= Tc2('E5',t);
inletTcE7('E7',t)..          Tc1('E7',t)    =e= Tc2('E6',t);
inletTcE8('E8',t)..          Tc1('E8',t)    =e= Tc2('E7',t);
inletTcE9('E9',t)..          Tc1('E9',t)    =e= Tc2('E8',t);
inletTcE10('E10',t)..        Tc1('E10',t)   =e= Tc2('E9',t);
inletTcE11('E11',t)..        Tc1('E11',t)   =e= Tc2('E9',t);
inletTcE12('E12',t)..        Tc1('E12',t)   =e= 0.56*Tc2('E10',t)+0.44*Tc2('E11',t);
inletTcE13('E13',t)..        Tc1('E13',t)   =e= Tc2('E12',t)-32;


inletThE1('E1',t)..          Th1('E1',t)   =e= 83;
inletThE2('E2',t)..          Th1('E2',t)   =e= Th2('E9',t);
inletThE3('E3',t)..          Th1('E3',t)   =e= Th2('E7',t);
inletThE4('E4',t)..          Th1('E4',t)   =e= Th2('E6',t);
inletThE5('E5',t)..          Th1('E5',t)   =e= 145;
inletThE6('E6',t)..          Th1('E6',t)   =e= Th2('E11',t);
inletThE7('E7',t)..          Th1('E7',t)   =e= Th2('E10',t);
inletThE8('E8',t)..          Th1('E8',t)   =e= 205;
inletThE9('E9',t)..          Th1('E9',t)   =e= 230;
inletThE10('E10',t)..        Th1('E10',t)  =e= 279;
inletThE11('E11',t)..        Th1('E11',t)  =e= Th2('E12',t);
inletThE12('E12',t)..        Th1('E12',t)  =e= Th2('E13',t);
inletThE13('E13',t)..        Th1('E13',t)  =e= 347;

FuelFIT(t)..                 FG(t)         =e=  -31.997*Tc2(‘E13’,t)+10139;
foulingy(i,t)..              Rf(i,t)        =e= (Rf(i,t-1)+dRf(i,t)-dRfhhx(i,t))*(1-h(i,t))*(1-
y(i,t))+(Rf(i,'0')*(h(i,t)+y(i,t)));
foulingUf(i,t)..             Uf(i,t)       =e= 1/((1/(Uc(i)+Uc(i)*sum(m,h(i,m))))+Rf(i,t));
outletTh(i,t)..              Th2(i,t)       =e= ((R(i)-1)*Th1(i,t) +(exp(((Uf(i,t)*A(i))
/(Fc(i)*Cc(i)))*(R(i)-1))- 1)*R(i)*Tc1(i,t))/(R(i)*exp(((Uf(i,t)*A(i))/(Fc(i)*Cc(i)))*(R(i) -1))-1);
outletTc(i,t)..              Tc2(i,t)          =e= Tc1(i,t)+(Th1(i,t)-Th2(i,t))/R(i);
cons1('E13',t)..             Tc2('E13',t)     =g= 200;
cons2(i,t)..                 h(i,t)+y(i,t)      =l= 1;
cons3(i)..                   sum(t,h(i,t+1)) =l= 0;
const4(i,'0')..              h(i,'0')             =l= 1;
CO2emission(t)..             CO2(t)             =e= 1.32*FG(t);

consE1('E1')..               sum(t,h('E1',t)+y('E1',t))=l= 1;
consE2('E2')..               sum(t,h('E2',t)+y('E2',t))=l= 1;
consE3('E3')..               sum(t,h('E3',t)+y('E3',t))=l= 1;
consE4('E4')..               sum(t,y('E4',t))=l= 60;
consE5('E5')..               sum(t,h('E5',t)+y('E5',t))=l= 1;
consE6('E6')..               sum(t,h('E6',t)+y('E6',t))=l= 1;
consE7('E7')..               sum(t,h('E7',t)+y('E7',t))=l= 1;
consE8('E8')..               sum(t,y('E8',t))=l= 60;
consE9('E9')..               sum(t,h('E9',t)+y('E9',t))=l= 1;
consE10('E10')..             sum(t,h('E10',t)+y('E10',t))=l= 1;
consE11('E11')..             sum(t,h('E11',t)+y('E11',t))=l= 1;
consE12('E12')..             sum(t,h('E12',t)+y('E12',t))=l= 1;
consE13('E13')..             sum(t,h('E13',t)+y('E13',t))=l= 1;

iRf(i,'0')..                 Rf(i,'0')               =e= 0;
```

```
dRfE1('E1',t)..         dRf('E1',t)      =e= 0.035;
dRfE2('E2',t)..         dRf('E2',t)      =e= 0.06;
dRfE3('E3',t)..         dRf('E3',t)      =e= 0.033;
dRfE4('E4',t)..         dRf('E4',t)      =e= 0.013;
dRfE5('E5',t)..         dRf('E5',t)      =e= 0.012;
dRfE6('E6',t)..         dRf('E6',t)      =e= 0.032;
dRfE7('E7',t)..         dRf('E7',t)      =e= 0.07;
dRfE8('E8',t)..         dRf('E8',t)      =e= 0.03;
dRfE9('E9',t)..         dRf('E9',t)      =e= 0.011;
dRfE10('E10',t)..       dRf('E10',t)     =e= 0.02;
dRfE11('E11',t)..       dRf('E11',t)     =e= 0.012;
dRfE12('E12',t)..       dRf('E12',t)     =e= 0.015;
dRfE13('E13',t)..       dRf('E13',t)     =e= 0.04;

foulinghex(i,t)..       dRfhhx(i,t)      =e= dRfhex(i)*h(i,'0');


cost..   z =e= sum(t,(FG(t)-FG('0')))*0.029*14.549*720 + sum((i,t),y(i,t))*40000;

Model scheduling /all/;
Solve scheduling using MINLP minimize z;
Display z.l,h.l,y.l;
```