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THE INFLUENCE OF FEED TEMPERATURE AND STRIPPERS' POSITION ON PRESSURE, TEMPERATURE, MASS FLOW AND CONTROL OF CRUDE DISTILLATION UNIT (CDU)

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ENDORSEMENT PAGE

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ABSTRACT

The crude distillation units are the first units used in oil refinery industry. From distillation units, petroleum fractions such as naphtha, kerosene, diesel and gas oil can be obtained for power supply and as feed in other industries. The aim of this paper is to get the feed temperature and strippers position which can provide highest profit in the refinery. Aspen Plus simulator was used to perform the simulation. The temperature of furnace on Preflash column and strippers position on the Pipestil column are the design variables. The quality of light ends, heavy naphtha and diesel in the 95% ASTDMD86 was set constant at 130°C, 190°C and 222°C respectively. The optimum condition was determined based on the maximum profit and it happened when the temperature in the furnace of preflash column was 345°C while strippers' position on pipestil was at S1p.5-4, S2p.14-13 and S3p.19-18 with a profit of USD54.93/m³. No significant changes was found on CDU profile when the feed temperature or stripper position was changed. The deviation that occurred was normal, therefore CDU operation was satisfying for all cases.

Keyword: Crude Distillation Unit (CDU), Stripper position, Optimization

PREFACE

All honor and glory be to the King Jesus Christ for His grace that made possible to start and finish the compilation of this report entitled “The influence of Feed temperature and Strippers’ Position on Temperature, pressure and Mass Flow in each Stage of Crude Distillation Unit CDU”. For the realization of this thesis the author also acknowledges some individuals who directly or indirectly provided their assistance, namely:

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Finally, I hope that this thesis report be useful now and in the future. The author recognizes there is deficiency in writing so suggestions and positive criticism from the readers are welcome.

Surabaya, August 07th 2016

The Author:

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LIST OF SYMBOLS

Symbol	Observation	Unit
CDU	Crude distillation unit	-
F_i^L	Feed mole flow of liquid from stage i	Kmol/hr
$F^v_{(i-1)}$	Feed mole flow of vapor from stage i	Kmol/hr
L_i	Liquid flow from stage i	Kg/hr
V_i	Vapor flow from stage i	Kg/hr
S_i	The mole flow stripper product from stage i	Kmol/hr
H_i	Enthalpy of vapor from stage i	J/kmol
h_i	Enthalpy of liquid from stage i	J/kmol
k_i	Vapor-liquid equilibrium ratio	-
H_L	Liquid hold up	m^3
A_{ai}	Active area of stage i	m^2
A_{di}	Downcomer area of stage i	m^2
F_L	Liquid rate	m^3/s
l_w	Weir length	m
CS	Chao-Seader	-
w	Acentric factor	-
$(h^o - h)$	Enthalpy departure or residual enthaply	J/kmol
$(h^o-h)^o$	Simple fluid term or first order enthalpy departure	J/kmol
$(h - h)^1$	The correction term or second order enthalpy departure	J/kmol
v^o	Liquid fugacity coefficient of pure component	-
$V^{(o)}$	Simple fugacity coefficient of fluid in the liquid state	-
$V^{(1)}$	Fugacity coefficient correction factor	-
Ø	Fugacity coeficient of the vapor phase	-
γ	Activity coefficient	-
δ	Solubility parameter	-

Symbol	Observation	Unit
δ_m	Solubility parameter for mixture	-
ΔE_v	Vaporization energy	J/kmol
Cs	Steam supply cost	US$/m^3
C_G	Total variable cost of raising steam	US$/Kg
V_{crude}	Volume of crude oil	m^3/hr
Ms	Mass of steam	Kg/hr
C_F	Steam fuel cost	US$/Kg
α_F	Fuel cost	US$/Kj
η_B	Overall boiler efficiency	-
Hs	Enthalpy of steam	Kj/kg
h_w	Enthalpy of boiler feedwater	Kj/Kg
C_{PH}	Cost of preheater heat	US$/m^3
E_{PH}	Energy required in preheater	Kj/hr
λ_S	Latent heat	Kj/Kg
C_{fi}	Furnace heat supply cost	US$/m^3
E_{fi}	Energy required in furnace	Kj/hr
B_f	Furnace fuel cost	US$/Kj
LHV_g	Fuel low heating value	Kj/kg
C_O	Blending crude oil purchase cost	US$/m^3
C_{Oi}	Individual crude oil purchase cost	US$/m^3
S_P	Total product sales	US$/m^3
S_{Pi}	Individual product sale	US$/m^3
K	Profit	US$/m^3
PF	Preflash furnace	

Symbol	Observation	Unit
PPF	Pipestil furnace	
LC	Level control	
LCW	Level control for water	
TC	Temperature control	
PC	Pressure control	
ITAE	Integral of the time –weighted absolute error	

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CHAPTER 1
INTRODUCTION

1.1 Background

Oil refinery is an industrial plant where crude oil is processed and refined into more useful products such as liquefied petroleum gases, petroleum naphtha, gasoline, diesel fuel, asphalt, heating oil and kerosene. Early refinery were predominantly distillation units, in 1861 kerosene, naphtha and tar was produced by atmospheric distillation process involving a simple batch distillation of crude oil with the objective of maximizing a kerosene production. In 1930 it was somehow more complex because cracking and cooking unit started to appear but was still essentially a distillation unit (Speight, 2011).

Nowdays the refinery is much more complex operation than the last 120 years, the process is normally based on distillation process using atmospheric or vacuum distillation column. Watkins (1979) described the crude fractions in three types - A, R and U which are characterized by their reflux, pump-back reflux, pump-around reflux and top reflux respectively. The product from distillation process goes to the conversion process where it will be resized and change their structure using various processes as thermal catalytic reforming which according to Simanszhenkov and Idem (2003) the temperatures of 510-565^{o}C are employed in the presence of hydrogen and moderate pressures to obtain gasoline with octane number of 70 to 80 from heavy naphtha having octane number of 40. In the alkylation process the benzene contained in the gasoline produced by thermal cracking process are removed and Muraza (2015) in his work removed the benzene from gasoline by alkylation using zeolite catalyst. The treatment process follows the conversion process and its aim is to prepare hydrocarbon fractions for further processing and transform it into the final product. The sulfur removing is an example, Yahaya et al. (2013) in their research removed sulfur compounds in diesel cut by using n-butyl-3-methyl pyridinium methylsufate as a supported ionic liquid membrane. Finally the formulation and mixing processes are carried out where fractionated hydrocarbons are mixture and some additives are added to obtain a determined product specification. Kirgina (2014) developed the recipe which used reformate, izomerizate, catalytic cracking gasoline, alkylate, isopentane and MTBE to obtain blended gasoline. Since the residue from this unit is still containing many solar fractions to maximize oil recovery an additional processing is carried out using a vacuum distillation unit (VDU). This is done because the residue consists of components with high boiling point, so that when it is processed through atmospheric distillation, most of the residue will undergo cracking due to the higher temperature required. Optimization in the operating conditions of CDU is required to achieve better economic benefits. In order to shift the production towards those distillates which carry added incentives for the refiners, optimization and instant rectification of the processing conditions are required. The quality of products needs to be analyzed by rigorous monitoring of the feed, Bagawejewcz (1997) studied about effect of pump-arounds and steam in CDU, Viswanathan (1993) found the optimum locations of trays feed for multiple feeds by maintain fix the entering reflux tray and entering location of the boilup, Seo et al. (2000) also in his work found an optimum feed tray with annual cost of $9.185.230 by considering the energy consumption, operating cost and annual cost as objective functions. Handogo (2011) obtained optimum feed temperature of 533K, the design variable was the temperature of preflash column the optimum condition was found by considering a maximum profit ($\$7.74/m^3$), Hossein, et al. (2013) studied about the liquid weeping and the hydrodynamic behavior of a column equipped with the circular sieve trays and Malvin (2014) developed a three-dimensional computational fluid dynamic to characterize flow regimes in a sieve by considering the pressure drop between trays and weep reflux.

The previous literature acknowledge that great progress has been achieved in this issue, especially the relationship between the feed temperature, energy recovery through the pump-around and the cost for all operational processes but still have lack of data of relation between these listed factor with the stripper position on CDU, which is the aim of this research. The design variable for this research is the feed temperature of preflash column and the strippers' position on pipestil column. CDU profile is examined in order to determine the condition that better influence it, by considering the 95% liquid volume ASTM D86 of products and profit.

1.2 Problem formulation

The central question of this research is about how the feed temperature and stripper position may influence the pressure, temperature and mass flow in each stage of crude distillation unit(CDU) maintaining the quality of 95% liquid volume ASTM D86 of. This research was chosen to increase the data of factor that influence the operation of CDU.

1.3 Research scope

This research has the following scope and limits:

1. The system under study is the CDU, from which the specified data of the feed will be extracted;
2. The simulation will be run in Aspen Plus V7.3 simulator;
3. The feed temperature and stripper position that better influence the pressure, temperature and mass flow in each stage of CDU will be found by comparing the advantages and objectives of this research.

1.4 Research Objective

The objective of this research is to determine the feed temperature and strippers' position which has a better influence on the pressure, temperature and mass flow of each stage into the crude distillation unit, in order to obtain a product according to the specification and with high profit.

1.5 Research benefits

This research will help to:

1. Understand the crude distillation unit(CDU);
2. Understand how the feed temperature and stripper position influence the pressure, temperature and mass flow in each stage into the CDU in order to obtain a product according to the specifications.

3. Get the feed temperature and stripper position that better influence the pressure, temperature and mass flow in each stage into the CDU in order to obtain a product according to the specifications.

CHAPTER 2
LITERATURE REVIEW

2.1 Optimization of CDU

The optimization tasks consist of finding the best solution for a given process, applying mathematical results and numerical techniques to the process in study. Ravindran et al. (2006) said that it is necessary to clearly delineate the boundaries of the system to be optimized, to define the quantitative criterion on the basis of witch candidates will be ranked to determine the best, to select the system variables that will be used to characterize or identify candidates, and to define a model that will express how the variables are related.

Optimization of crude CDU becomes more and more important case of study due to the ecological requirements and high energy costs. Optimization of CDU has been presented by various academic contributions especially in CO_2 remover and heat exchanger network and most of them were using computer simulaion. Gandalla (2013) in his work, optimization based on retrofit approach for revamping an Egyptian CDU was performed by changing the column operating conditions like pumparound flow rates, reflux ratio, stripping steam rate and temperature across the pumparounds to reduce the energy consumption. Oni (2014) improved a performance of an existing CDU increasing 4% of its initial overall energy by using process simulation techniques and combined exergy and retrofit methods to show what the process is capable to achieve under considerable expense on the required capital investment. Recently, Luo (2015) in his paper shows a systematic optimization approach produces a maximum annual economic benefit of an existing crude oil distillation system by considering product output value and energy consumption simultaneously.

2.2 Description of crude oil distillation process in CDU

In the processing of crude oil, CDU is the first unit and it aims to separate the mixture into various fractions such as naphtha, kerosene, diesel, atmospheric gas oil and residue depending on their boiling points and each fraction can then be moved to other refinery units. CDU is known as an atmospheric distillation unit and it operates at pressures slightly above atmospheric pressure.

Table 2.1 CDU product

Cut	Product	End point (°C)
1	Off gas	10
2	Light straight run naphtha	70
3	Naphtha	180
4	Kerosene	240
5	Light diesel	290
6	Heavy diesel	340
7	Atm. Gas oil	370
8	Vacuum gas oil	390
9	Vacuum distillate	550
10	Vacuum residue	-

Source: Fahim et al., 2010

Crude oil contains dissolved salts as a piece of composition and to prevent equipment damage, the oil is desalted and liquefied before heading to the CDU. Within the desalter, the crude oil is mixed with water in order to dissolve the salt contained in the crude oil, then the salty water is separated and the amount of water which is left in the crude oil is called diluted water. The desalted crude oil is preheated in a heat exchanger and follows to the furnace where it is reheated to about 340-372°C to promote a partial evaporation before entering in the main column.

The column usually has 25 to 35 stages, 2 or 3 pumparounds which have the function to ensure the reflux inside the column and Oh (2000) said that pumparounds is one of the ways to change temperature profile. It has also one condenser and three side streams for kerosene, diesel and atmospheric gas oil productions. To obtain a clear separation, strippers are coupled on the main column in every side stream, the strippers are fed with steam from the bottom in order to vaporize lifted light components. By using steam to stripping the main column and stripper columns an inexpensive separation process can be achieved, note that reboilers are avoided into the CDU. Bagajewics (1997) also said that the usage of reboiler in the main fractionation column and side-strippers as opposed to the steam injection has been ruled out due to the high temperatures needed for a certain degree of separation. After partial vaporization, the oil goes into the pipestil flash zone and produces heavy naphtha as distilled, residue as bottom product and kerosene, diesel and atmospheric gas oil as side-stream product (Figure 2.1).

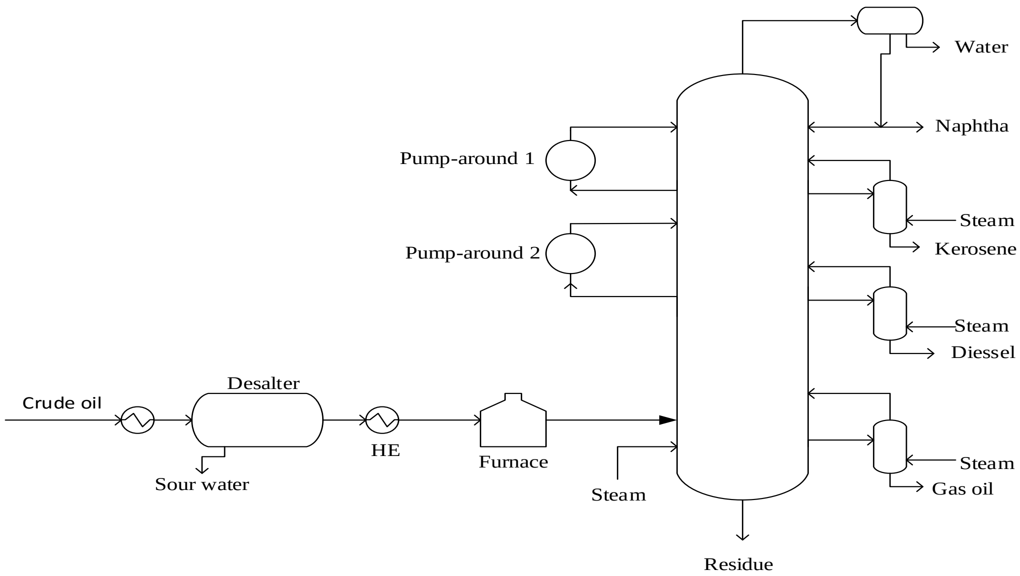

Figure 2.1 Pipestil column configuration according to Bagajewics (1997)

2.2.1. The striping type design

In the stripping type design, the crude is heated about 160°C and fed close to the top of column, since the crude temperature is low, the vapor ratio of the feed is small. The crude goes down the column and is heated consecutively with the upper, middle and lower heater (Figure2.2). Unlike conventional design, in this design the light components are withdrawn from top section as soon as they are vaporized, they do not reach the trays where diesel and gas oil are witdrawn.

Ji and Bagajewicz (2002) said that the stripping type design cannot achieve the same low yield of residue as the convenctional design, which makes an energy efficiency comparison pointless and if the maximum temperature limit is increased over the cooking limits the stripping type design and the energy efficience matched decrease the 5% improvement over the conventional design, so that the stripping type is not competitive.

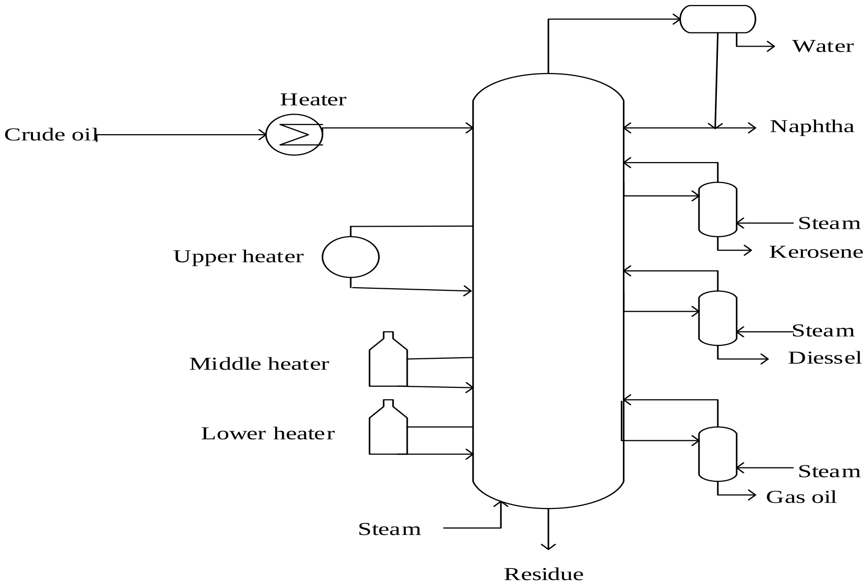

Figure 2.2 Stripping type column from Ji and Bagajewics, 2002

2.2.2. The installation of preflash drum

Preflash drum unit are used when light crude oils are being processed because the pressure required to suppress vaporization is high. In this case, the best thing to do is separate some lights components before heating the crude further in the preheat train, consequently it will reduce the operating pressure of the main furnace of the CDU. A drum is a single-stage flash that separates vapor from the liquid feed and it must retain foam that is created. So, the amount of hydrocarbons vaporized depends on the temperature and pressure in the drum, the flashed crudes leaves the bottom of the drum and the foam-free vapor stream should exit the top and are sent directly to the atmospheric column. Since the preflash drum is undersized and do not retain the foam, the preflash drum vapor is fed to the main column where the composition matches the endpoint of the drum vapor as shown in figure 2.3.

Al-Mayahi (2014) in his work were investigate the effect of preflash drum designs on energy efficiency associated CO_2 emissions of the CDU, showed that the introduction of crude pre-flash has another vantage which is a noticeable reduction in the CO_2 emissions of the CDU.

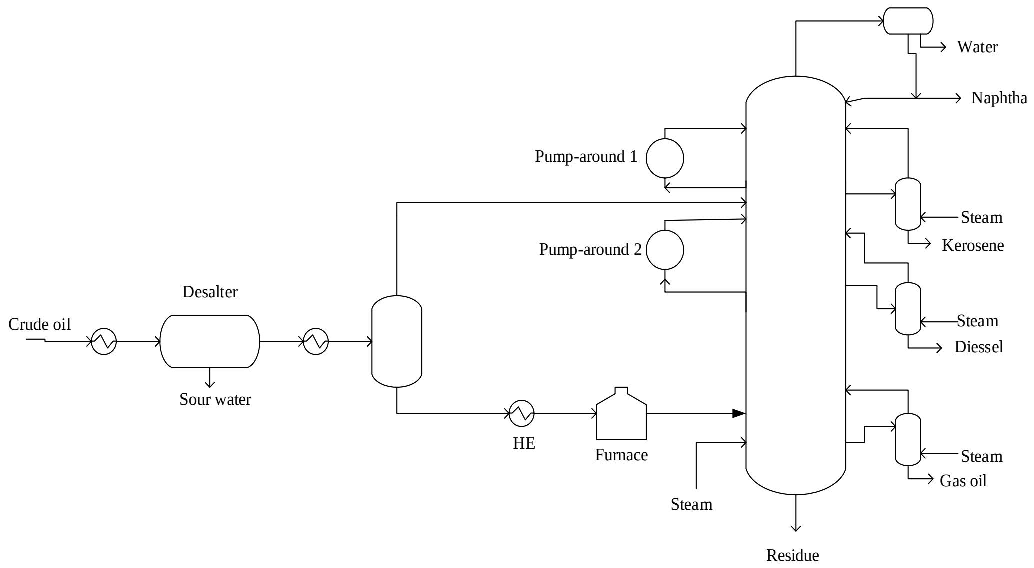

Figure 2.3 Preflash drum and Pipestil configuration from Al-Mayahi, 2014

2.2.3. The installation of preflash column

Pre-fractionation columns use trays and reflux to fractionate overhead product from side-draw or bottom product streams. Unlike preflash drum where the components in the light naphtha range are condensed in the condenser of the main column, this column has their own overhead condenser where those components in the light naphtha range are condensed in, and heavy naphtha are condensed in the condenser of the main tower.

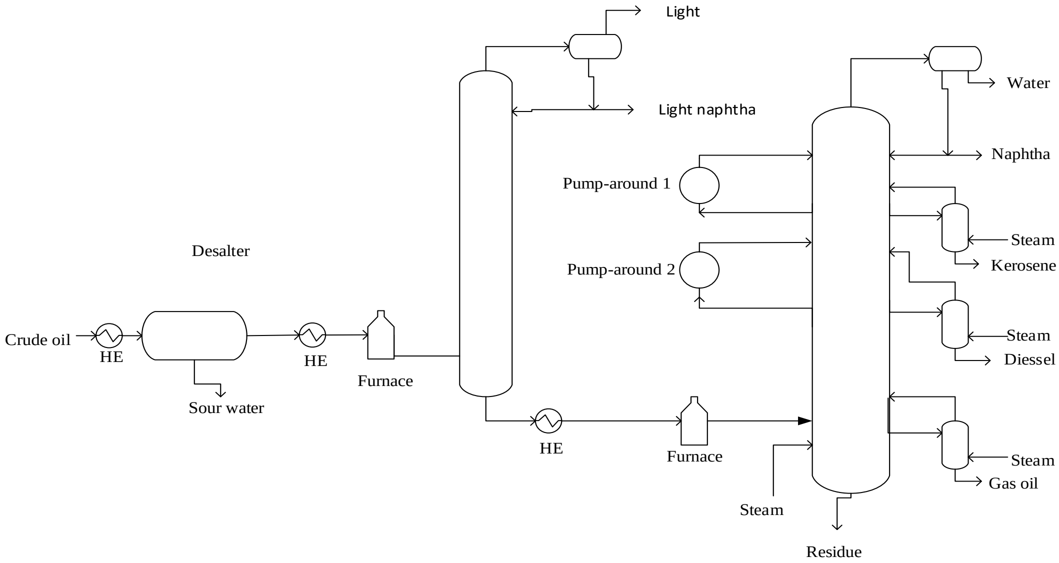

Figure 2.4 Preflash column and Pipestil configuration

The residue from Pipestil contain a significant amount of valuables oil that cannot be distillated at atmospheric pressure because the temperature needed to proceed the separation is so high and thermal cracking can takes place. Pujado & David (2006) said that at atmospheric condition to achieve any meaningful degree of vaporization the flash zone temperature would be extremely high (in excess of 482°C) to break or cracking. Wauquier (1998) stated that the reduce oil is heated in a furnace at a maximum temperature of some 380 to 415°C or from 365 to 400°C at the column inlet after isenthalpic flash along the transfer line, and fed into the vacuum distillation column.

2.2.4. Variables of the CDU

Flash zone temperature - The temperature of the flash zone is set so that a certain required amount of feed is flashed, and the higher the temperature, the more top product and the lower bottom product, but this should not be greater than the allowed temperature because it can cause thermal decomposition (cracking) of the oil.

Top column pressure - The pressure at the top is fixed by the designer, selected based on the available average condensation temperature in the condenser and there will be also the difference between the bottom pressure and the trays column pressure drop. The pressure depends on vapor pressure of the liquid at the bottom of the column, which in turn also depends on the temperature of the bottom column, if the temperature increases the vapor pressure of the liquid will increase and consequently the column pressure and the temperature for each tray will increase.

Top temperature column - The top temperature is one of the indicators of naphtha end point, this should be high enough so that the top product will evaporate completely. If the temperature is higher, so many heavy components (unwanted) also will be vaporized and will be in the top product, but if the temperature is lower certain amount of desired product is condensed. If there are product side draw, those amount of condensed product will be incorporated.

According to Fahim (2010), the top temperature must be controlled to be 14 - 17^{o}C higher than the dew point temperature for the water at the column overhead pressure so that at liquid water is condensed, this is to prevent corrosion due to the hydrogen chloride dissolved in liquid water.

Stripping stream - In the CDU, stripping steam is used to remove lights ends embedded in side-draw products, in order to obtain product with a maximum purity in an inexpensive separation process that is why the use of reboilers in stripper is replaced by steam.

The steam when it comes in contact with the oil fraction, reduces their partial pressure of vaporization, what will allow the unwanted dissolved hydrocarbons be vaporized at low temperatures. The steam can also be introduced starting from the bottom of the main column with the same purpose as mentioned above which is removing light components dissolved in heavy petroleum fractions.

2.2.5. Mathematical modeling

As for binary systems, the calculation of equilibrium stages of a multicomponent distillation process is performed by the normal procedure which is the resolution of the MESH equations step by step. Mass balance, energy and vapor-liquid equilibrium can be written for each component or pseudo-components present in the mixture as well as an overall balance for the entire column or for each stage. But as before reported by Saraf (2001) said that from a practical point of view, it is not possible to represent the feed crude oil distillation or its products in terms of current component flow rates or mole fractions since crude oil is a mixture of several hundred constituents that are not easy to analyze. The Generally Accepted practice is to express composition of crude oil in terms of a finite number of pseudo-components. Each pseudo-component is treated as a single component, is in fact a complex mixture of hydrocarbons with a range of boiling points within the narrow region say 25 °C wide.

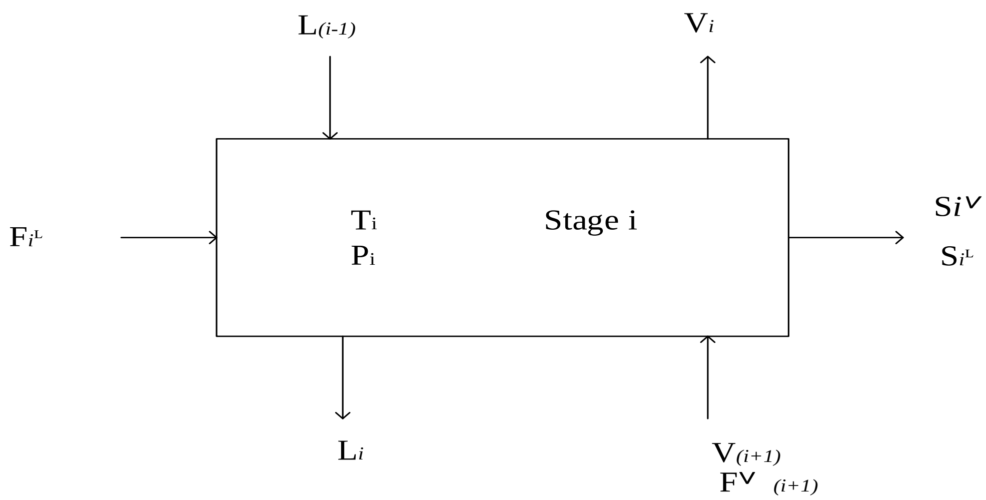

Figure 2.5 Scheme of a column stage

Mass balance

$$L_{(i-1)} + V_{(i+1)} + F_i^L + F_{(i+1)}^V = L_i + V_i + S_i \tag{1}$$

Component balance

$$L_{(i-1)}x_{(i-1)j} + V_{(i+1)}y_{(i+1)j} + F_i^L x_{ij} + F_{(i-1)}^V y_{ij} = L_i x_{ij} + V_i y_{ij} + S_i x_{ij} \tag{2}$$

Energy Balance

$$L_{(i-1)}h_{(i-1)} + V_{(i+1)}H_{(i-1)} + F_i^L h_i + F_{(i-1)}^V H_i = L_i h_i + V_i H_i + S_i h_i \tag{3}$$

Thermal properties equation

$h_i = (x_{ij} T_i)$

$H_i = f(y_{ij,} T_i, P_i)$

$h_i^F = f(x_{ij}^F, T_i^F)$

$H_i^F = f(y_{ij}^F, T_i^F, P_i)$

Vapor-liquid equilibrium

As in the case of binary mixtures, the vapor-liquid equilibrium for a multicomponent mixture is described by using the distribution coefficients or factors K. Each component has its K factor, which is defined as the ratio of the component fractions present in the vapor phase and liquid phase at equilibrium.

$$K_{i,j} = {}^{y_{i,j}}\!/_{x_{i,j}}.\ [i=1,...,N;j=1,...,C] \tag{4}$$

$$\sum x_{i,j} = \sum y_{i,j} = 1 \tag{5}$$

For cases of total condenser, the equation is written as follows:

$$x_{i,j} = {}^{l_{i,j}}\!/_{L_{i,j}} \qquad [I = 2, ..., N] \tag{6}$$

$$K_{i,j} = {}^{l_{i,j}}\!/_{L_{i,j}} \tag{7}$$

For a given stage, the liquid hold up (H_L) can be determined using the following equation:

$$H_L = h_{ai}\rho_{Li}A_{ai} + h_{di}\rho_{Li}A_{di} \tag{8}$$

Where:

h_{ai} and h_{di} is the net height of the stage and downcomer;

ρL_i is the liquid density in the stage;

A_{ai} and A_{di}, are the active area of the stage and the downcomer.

The Francis equation can be used to relate H_L at a certain stage and and liquid flow.

$$F_L = 3.33 * 0.5286 * l_w * \sqrt[3]{h} \tag{9}$$

Where F_L is liquid flow rate, the length l_w and weir length h of the liquid exiting the weir or weir crest.

2.3 Software Aspen Plus

Aspen Tech is a provider of software and services for the process industries, and Aspen Plus software is one of them, which is simple software that facilitates the modeling of processes in oil refineries. Such processes are complex and interrelated, this is the unique feature that makes them different from other processes, starting with their own feed which is a complex mixture of hydrocarbons with a wide difference in boiling point.

Through Aspen Plus can be developed models of simulation processes to processing crude oil and then use that model to find better alternatives of operation or simply optimize already existing processes. In oil refinery this units of separation are interlinked, every unit having its specific product.

Usually the preflash column is the first in the sequence, used when the mixture contains lights component or to unload the atmospheric furnace, to eliminate vaporization at the furnace inlet control valves, to increase the naphtha production and to debottleneck the crude column system. Luyben (2006) says that in Aspen plus the Preflash tower is simulated as a Preflash column and Errico (2008) in his work said that naphtha reduction in the top of the main column is an aspect to be considered. So, if we want to keep the same end point for the naphtha stream, the top temperature decrease with possible condensation phenomena and consequent corrosion Possibility, in his work they used top temperature value (higher than 100^{o}C) to avoid corrosion and a long running time apparatus.

The fleshad liquid is then pumped into the Pipestil, the pumps as well as valves are standards. The column which are used are Petroleum fractionator type, chosen in the library menu model that found at the bottom of the Aspen plus window, one click on Petrofrac column and various options appear.

2.3.1 Thermodynamic model

Various thermodynamic packages designed to light hydrocarbons, and gases such as Braun K-10 packets, Grayson-Stread and Chao-Seader (CS) are used in the Aspen Plus simulator. Aspen Plus (2006) states that the Braun K-10 package is suitable for processes involving heavy oil and low pressure fractions, best results will be obtained when using mixture of pure aliphatic or aromatic as wel as mixtures involving light gases and high pressures the CS and Grayson-Stread packages are recommended.

Aspen Plus recommends using CS package, which is widely used in research papers as the case of Haydary & Pavlik (2009) research used for simulation in steady-state and dynamic of CDU and Handogo (2010) to optimization the CDU. However Gutierrez et al (2014) said that although the theorical base of CS package is not solid, simulator compensates it with their wide crude database from all around the world. Its application has to be avoided is outside its valid range, Aspen Tutorial recommends this package because its application is reliable thanks to their wide source of information. The CS package uses the following sub-packages:

Lee-Kesler correlation for calculation of enthalpy

The proprieties of a real fluid in the Lee-Kesler method are related to be properties of a simple fluid (w = 0) and those of reference fluid. The basic parameters of the model are reduced pressure and temperature which should be calculated based on mixing rules.

$$\left(\frac{h^o-h}{RT_c}\right) = \left(\frac{h^o-h}{RT_c}\right)^o + w\left(\frac{h^o-h}{RT_c}\right)^1 \qquad (10)$$

$(h^0 - h)$, $(h^0 - h)^0$ and $(h - h)^1$ are the enthalpy departure or residual enthalpy, simple fluid term or first order enthalpy departure and the correction term or second order enthalpy departure. YVC Rao (2004) said that

$\left(\frac{h^o-h}{RT_c}\right)^o$ *and* $\left(\frac{h^o-h}{RT_c}\right)^1$ are Evaluated in the temperature range Tr = 0.3 to 4.0 and pressure range Pr = 0.01 to 10.

Liquid fugacity coefficient of pure component (v $^{\mathbf{0}}$)

The liquid fugacity coefficient of pure component is a well-defined property when the component exists as a liquid, but at condition where does not exists a pure liquid, this value is hypothetical, such conditions occur in mixtures.

$$\log v^0 = \log v^{(0)} - w \log v^{(1)} \quad (11)$$

$$\log v^{(0)} = A_0 + \frac{A_1}{T_r} + A_2 * T_r + A_3 * T_r^2 + A_4 * T_r^3 + P_r(A_5 + A_{6*}T_r + A_7 * T_r^2) + P_r(A_8 + A_{9*}T_r)$$

$$\log v^{(1)} = -4.23893 + 8.65808 * T_r - \frac{1.2206}{T_r} - 3.15224 * T_r - 0.025 * (P_r - 0.6)$$

v $^{(o)}$, v $^{(1)}$ are simple fugacity coefficient of fluid in the liquid state and fugacity coefficient correction factor.

Redlich-Kwuong state equation for fugacity coeficient of the vapor phase (∅)

$$ln\emptyset_i = ln\left(\frac{v}{v-b}\right) + \frac{b_i}{v-b} - \frac{2\sum y_i a_{ij}}{RTb} ln\left(\frac{v}{v-b}\right) + \frac{ab_i}{RTb^2}\left(ln\frac{v}{v+b} - \frac{b}{v+b}\right) - lnz \quad (12)$$

Scatchard-Hildebrand model for activity coefficient (γ)

The Hildebrand's equation is expressed in terms of properties of pure components and the solubility parameter (δ) is the square root of the ratio of the vaporization energy (ΔEv) and molal volume of liquid (V).

$$\delta_i = \left(\frac{\Delta E_v}{V}\right)^{1/2} \quad (13)$$

Chao & Seader (1961) said 'as a matter of pratical convenience the standard reference temperature was chosen to be 25°C, and the extensive tabulation of density and heat of vaporization of a large number of hydrocarbons at this temperature in API Project 44 were then immediately available for the calculation of V and S'.

The solubility parameter of the blend is given by the volumetric average of the components. According Albright (2009) those parameters are available for numerous substances, and additional parameters can be readily determined from their definitions, when the need arise, except for light gases.

$$\delta_m = \frac{\sum x_i V_i \delta_i}{\sum x_i V_i} \tag{14}$$

$$ln\gamma_i = \frac{V_i * (\delta_i - \delta)^2}{RT} \tag{15}$$

The equilibrium value is then calculated using the formula:

$$K_i = \frac{v_i \gamma_i}{\emptyset_i} \tag{16}$$

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CHAPTER 3
RESEARCH METHODOLOGY

3.1 Research stages

From this research, is expected to obtain the feed temperature and stripper position that better influence the pressure, temperature and mass flow in each stage of the Preflash and Pipestil column. The figure 3.1 shows the methodological sequence that will be used to reach this objective.

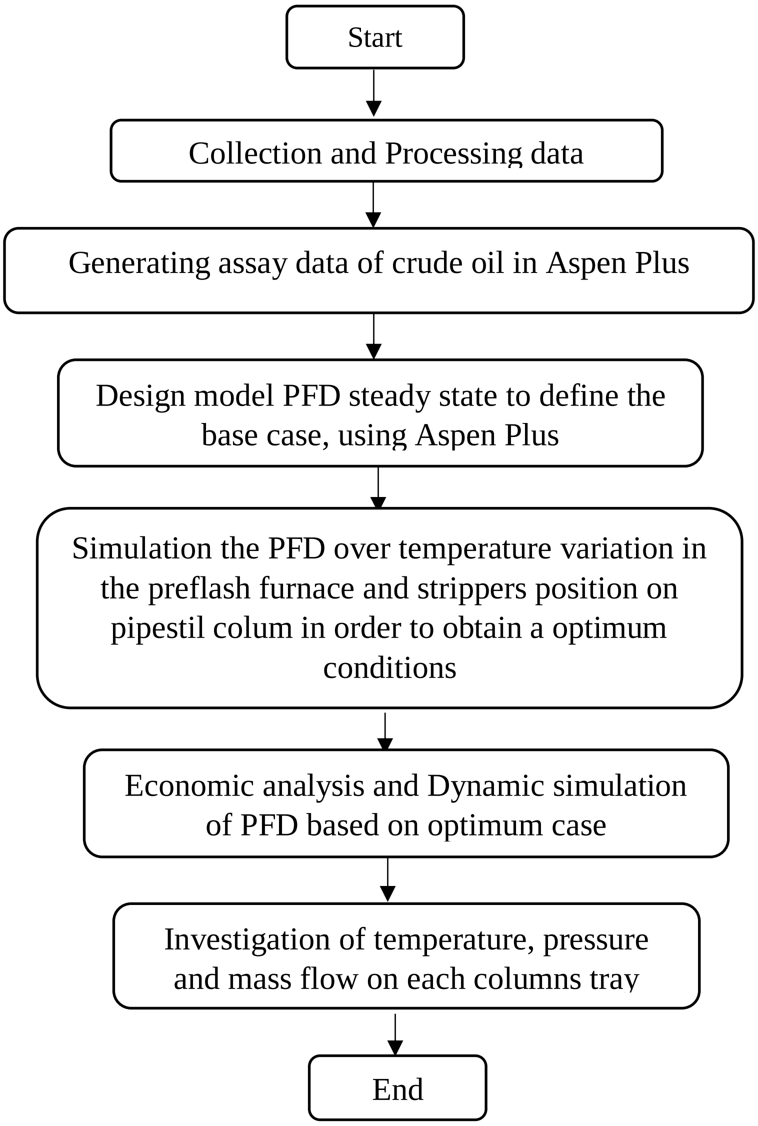

Figure 3.1 Research flow diagram

3.2 Collection and Processing data

For this research are necessaries the following essay data of crude oil:

Table 3.1 Assay data of crude oil (40.57 °API)

TBP (°C)	Distillation volume (vol.%)	Light-ends components	Analysis (Liquid volume fraction)
35	0	Isobutane	0.002946
100	5.7	n-Butane	0.005301
113	10	Isopentane	0.01132
142	20	n-pentane	0.002946
150	22	Hexane	0.02983
169	30		
194	40		
200	42		
217	50		
245	60		
250	62.3		
273	70		
300	76.7		

Source: Puspitasary and Setyarini, 2010

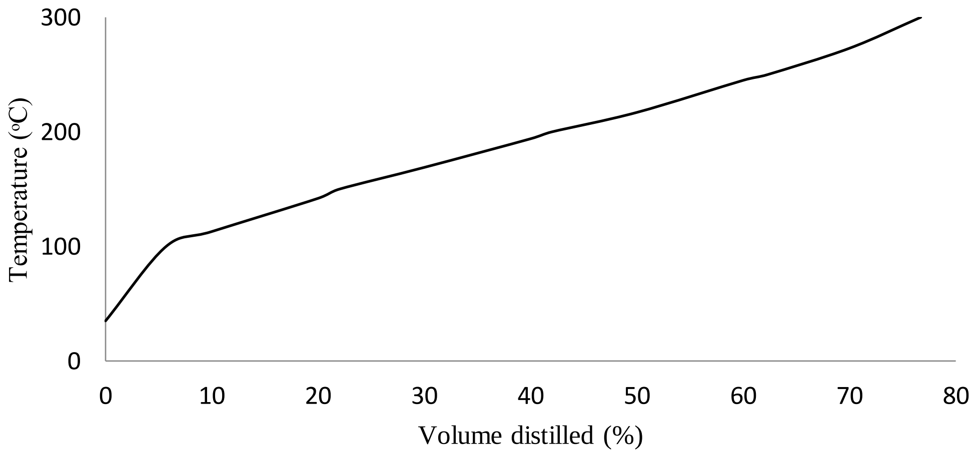

Figure 3.2 TBP curve from essay data of blended crude oil

Table 3.2 Estimated initial composition of blended crude oil

Products	Boiling range (°C)	Acc. volume distillated (%)
Light	IBP < 40	0.42
Light naphtha	40 - 125	5.8
Heavy naphtha	90 - 150	10.2
Kerosene	113 - 169	16.34
Diesel	140 - 217	21.64
AGO	169 - 378	52.15
Residue	355 < FBP	100

From figure 3.2 was estimate the initial composition of oil, from it can be said that from total crude oil exist only 0.42% of light ends and the largest amount portion is the AGO about 30.51%. These data reveal that this blended oil belongs to the naphthenic group (Simanzhenkov and Idem, 2003).

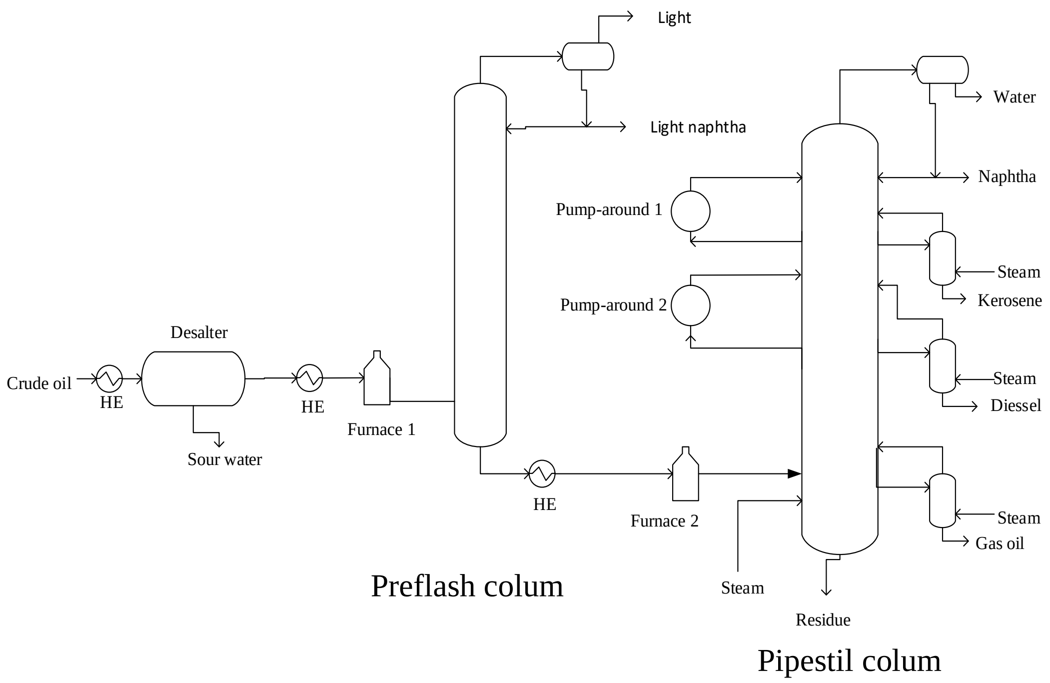

Figure 3.3 Crude distillation units according to Luyben, 2013

3.3 Research variables

3.3.1 Design variable

The design variable for this research is the temperature of preflash furnace and strippers position on pipestil column, the temperature shall be vary from 260 °C to

350^{o}C to avoid the cracking of light fraction, but during the simulation the pumparround flow rate will be fixed.

3.3.3 Variables to be observed in each stage and in the output stream

The variables to be observed in each stage are the temperature, pressure and both mass of liquid and vapor leaving the stage while in the output stream are: stream product, reflux ratio and furnace heat duty of each unit.

3.4 Generating data assay crude oil in Aspen Plus

The generation of pseudo-components is the first step on simulation, since the crude oil is mainly composed by complex mixture of hydrocarbons, so that will be better generate a pseudo-component. Aspen Plus simulator has a special run to generate it.

3.5 Design model PFD steady state to define the base case, using Aspen Plus

The simulation will be performed in steady state, the first simulation will be used as base case of the optimization step. The simulations will be run following example from Luyben (2013), but with various alterations as showed in table 3.2 and 3.3.

Table 3.2 Design variable for the base case simulation

Column	Variable	Value
	Main feed (Kg/hr)	85000
Preflash	Preflash main steam, Kg/hr (4.14 bar 300^{o}C)	790
	Preflash condenser temperature (^{o}C)	72
	Preflash tray efficiency (%)	60
	Pipestil main a steam, Kg/hr (4.14 bar 300^{o}C)	1260
	Pipestil tray efficiency (%)	60
	Pipestil condenser temperature (^{o}C)	91
	Overflash (%)	3
	Kerosene stripper steam, Kg/hr (4.14 bar 300^{o}C)	324
Pipestil	Diesel stripper steam, Kg/hr (4.14 bar 300^{o}C)	90
	AGO stripper steam, Kg/hr (4.14 bar 300^{o}C)	72
	Light naphtha D86 95% temperature (^{o}C)	138
	Heavy naphtha D86 95% temperature (^{o}C)	190
	Diesel D86 95% temperature (^{o}C)	222
	Pump-around 1 (PA1) return temperature (^{o}C)	138
	Pump-around 2 (PA2) return temperature (^{o}C)	186

Table 3.3 Base case column characteristics

Column	Variable	Value
Preflash	Number of tray in preflash column	10
	Preflash feed tray	10
	Tray spacing in Preflash (m)	0.61
Pipestil	Number of tray in Pipestil column	25
	Pipestil feed tray	22
	Tray spacing in Pipestil (m)	0.61
	Pump-around 1 (PA1) draw tray	8
	Pump-around 1 (PA1) return tray	6
	Pump-around 2 (PA2) draw tray	14
	Pump-around 2 (PA2) return tray	13
	Kerosene stripper draw tray	6
	Kerosene side stripper return tray	5
	Number of tray in kerosene stripper	4
	Diesel stripper draw	14
	Diesel side stripper return tray	13
	Number of tray in diesel stripper	3
	AGO stripper draw tray	19
	AGO side stripper return tray	18
	Number of tray in AGO stripper	3

3.6 Simulation the PFD over temperature variation in the preflash furnace and strippers position

At this moment, the position of stripper on pipestil column and furnace temperature on preflash column will be varied but the remaining data used until this phase will be the same as the previous stage.

3.7 Economic analysis

1. Steam supply cost

$$C_S = \frac{M_S * C_G}{V_{crude}} \tag{17}$$

The total variable cost of raising steam (C_G) normally is accounting as a much of 90% of the total cost (C_F), which is given by:

$$C_F = \frac{\propto_F * \eta_B * (H_s - h_w)}{1000} \tag{18}$$

2. Preheater supply cost

$$C_{PH} = \frac{E_{PH} * M_s * C_G}{\lambda_s * V_{crude}} \tag{19}$$

3. Furnace supply cost

$$C_{fi} = \frac{E_{fi} * B_f}{LHV_g * V_{crude}} \tag{20}$$

4. Purchase cost of blending crude oil

$$C_O = \frac{\sum x_i V_{ci} C_{Oi}}{V_{crude}} \tag{21}$$

5. Value of product sales

$$S_p = \sum S_{pi} * V_i \tag{22}$$

6. Profit calculation

$$K = S_p - (C_S + C_{PH} + \sum C_{fi} + C_o) \tag{23}$$

After getting result from Aspen plus, the profit is calculated to see how the variation of feed temperature on Prelash column temperature and strippers positions affect it, since all conditions of operation have been tried in attempt to produce more diesel and kerosene which are expensive and higher demand products. The case which have the higher profit will be considered as optimum case. The profit calculations can be explored more in Handogo (2011) and Pupistasari and Setyarini (2010).

3.8 Optimization

Optimization is to obtain the potential of this research that is determining the feed temperature that best influence the temperature, pressure and mass flow of each stage in CDU. In this stage, the profit obtained from these cases will be compared. The case at which will found the highest profit will be the focus of analysis of the infuence of feed temperature and strippers position.

3.9 Dynamic simulation of PFD based on optimum case

Since the steady state simulation and economic analysis is done, dynamic simulation was performed to observe the response of products composition to a change of the feed flow rate, the controlled variable was the temperature of 95% of the ASTM D86 curve representing the composition. The pressure units such as valves and pumps, that are not necessary for the steady-state simulation was specified for this simulation. Sizing of the equipment is another requirement of the dynamic simulation. The column diameter, tray spacing, tray active area, weir length, weir height, reflux drum length and diameter are requested for the dynamic simulation. All control charectesic of flow, temperature, pressure and level are shown in table 3.4, 3.5, 3.6 and 3.7 respectively, all of them are based on Luyben (2013)

Table 3.4 Flow control characteristics

Description	FCmixoil	FCpf-stm	FCpp-stm	FCstg19
Gain	0.5	0.5	0.5	0.5
Ţi(min)	0.3	0.3	0.3	0.3
SP (m^3/h)	713	2268	5444	68939
PV (Kg/h)	0 – 1425	0 - 4536	0 - 10888	30000 - 90000
OP (%)	50	50	50	50
Action	Reverse	Reverse	Reverse	Direct

Table 3.5 Temperature control characteristics

Description	TCfurnace1	TCfurnace2
Gain	0,63	0,63
Ʈi(min)	5,28	5,28
Dead time (min)	1	1
SP (^{o}C)	232	356
PV range (^{o}C)	0 - 464	0 - 713
OP	50	50
Action	Reverse	Reverse

Table 3.6 Pressure control characteristics

Description	PCcond1	PCcond2	PC-S1	PC-S2	PC-S3
Gain (%)	20	20	20	20	20
Ʈi (min)	12	12	12	12	12
SP (bar)	2.3	1.08	1.46	1.55	1.6
PV range (bar)	1.4 – 4.7	0.5 – 2.6	0.6 – 3	0.6 – 3	0.5 – 3.5
OP (%)	50	50	50	50	50
Action	Direct	Reverse	Direct	Direct	Direct

Table 3.7 Level control characteristics

Description	LC1	LC2	LC-W1	LC-W2	LCcond1	LCcond2
Gain	2	2	2	2	2	2
Ʈi(min)	9999	9999	9999	9999	9999	9999
SP (*m*)	2.4	0.068	0.025	0.02	0.79	0.42
PV range (*m*)	0 - 4.8	0 – 0.14	0 – 0.5	0-0.039	0 – 1.6	0 – 0.84
OP (%)	50	50	50	50	50	50
Action	Direct	Direct	Direct	Direct	Direct	Direct

3.10 Investigation of temperature, pressure and mass flow on each columns tray

In this step, the temperature, pressure and mass flow of each stage will be listed and compared with a theory to analyze the influence of feed temperature strippers' position on them.

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CHAPTER 4
RESULTS AND DISCUSSION

4.1 Generating essay data of crude oil in Aspen Plus simulator

The represention of crude oil and their cuts is not pratical in terms of components because they are complex mixtures and then it is necessary to represent them in terms of small fractions or pseudocomponent based on the distillation curve. Before designing the model PFD steady state, assay data for blended oil was made using Aspen Plus based on its laboratorium data (see table 3.1). This simulation was perfomed in order to predict the oil composition and get a complete percentage range (0% to 100%) of distilled oil and correspondent boiling point as expected, while in the laboratory data the range is 0% until 76.7% only. After data has been generated, both laboratory data and Aspen Plus data were used to plot a TBP curve in order to compare predictions as shown in figure 4.1 and table 4.1.

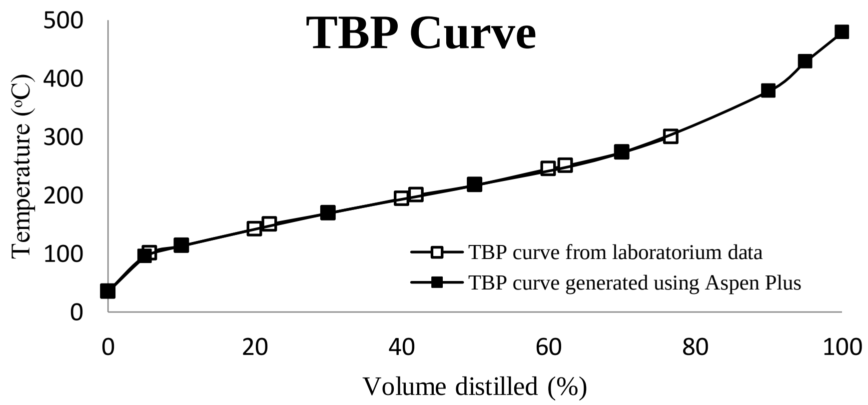


Figure 4.1 Comparison of TBP curve from laboratorium data with that generated from using Aspen Plus

From this graph the initial composition of the oil cuts is estimated, the composition is given in percentage of distilled volume as showed in table 4.1. In this table it is observed that the light components predicted is the smallest, about 0.42 and 0.62% for laboratoty and Aspen plus data respectivelly.

The greater prediction cover the AGO cuts; 30.51% and 33% for laboratoty and Aspen plus respectivelly. Although there is an insignificant deviation between the prediction made through laboratory data and aspen Plus, the results are satisfying and show how they are related.

Table 4.1 Estimated initial composition of crude oil

Products	Boiling range (^{o}C)	Volume distillated (%) laboratory	Volume distillated (%) Aspen plus
Light	IBP < 40	0.42	0.62
Light naphtha	40 - 125	0.58	9.52
Heavy naphtha	90 - 150	10.2	12.32
Kerosene	113 - 169	16.32	18.42
Diesel	140 - 217	21.64	21.22
AGO	169 - 378	52.15	54.22
Residue	355 < FBP	100	100

From the total crude oil used only 1.2% of light components was obtained and the largest amount was registered in AGO 48% . These data reveal that the crude oil used in this simulation belongs to the naphthenic group (Simanzhenkov and Idem, 2003).

4.2 Design model PFD steady state to define the base case, using Aspen Plus

In this research a Preflash column having trays and top reflux which corresponds to the conventional top condenser of distillation column were used. This pre-fractionator was conected to the main column called pipestil. The main colum has pump-arround reflux, top reflux and strippers as shown in figure 4.2. The simulation was made in Aspen Plus, the chosen thermodynamic package was CHAO-SEADER because crude oil is a complex mixture of hydrocarbons and it is impratical to be specified, this package is suitable for this kind of mixture (Hayday and Pavlik, 2009).

For defining the base case, the temperature in preflash furnace was set to be 260^{o}C and the positions of these three strippers on pipestitil were S1p.6-5, S2p.14-13 and S3p.19-18 respectivelly. The subscript S and the number that precedes it indicate the stripper and its number, the subscript “p” and the number before the dash indicates the tray at which the liquid draws from the main column to the strippers while the number after the dash indicates the tray at which the top product of strippers back to the main column. The input and the results from this simulation are shown below. When the simulation over temperature variation in the preflash furnace was carried out, other variables that influence the distillation on CDU such as the temperature on the top of column, the stripping steam flow rate, and the pressure on the top of column was maintained constat only varies the flash zone temperature of column throught the variation of furnace temperature on preflash column. Since the overflash and operation pressure was set to be 3% and 3.3 bar respectively, thus the temperature in the pipestil furnace will be dependent on the temperature of the residue coming from the Preflash column, so the higher the residue temperature from preflash column the lower will be the heat duty required on pipestil furnace (Puspitasary and Setyarini, 2010). This fact is clearly show on the table bellow.

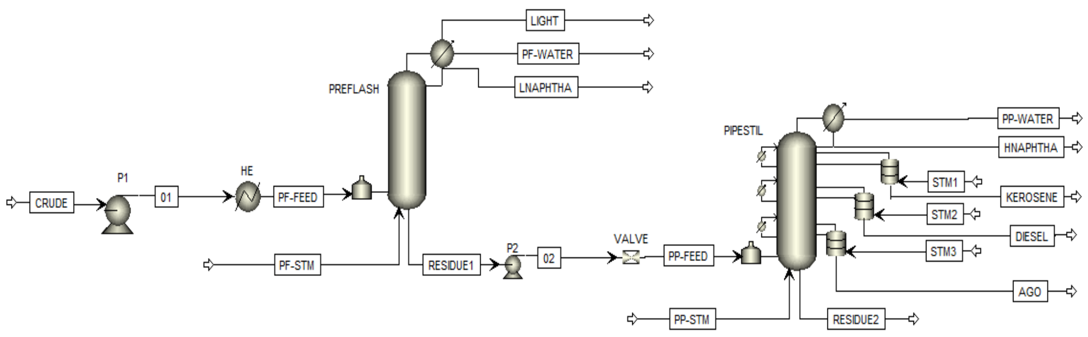


Figure 4.2 PFD of the base case simulation according to Luyben, 2013

Table 4.2 Result from steady state simulation of base case

Column	Description	Temperature of preflash furnace 260(°C)
Preflash	**Input (Kg/hr)**	
	Crude oil	85000
	PF-STM	790
	Total input	85790
	Product (Kg/hr)	
	LIGHT	983
	LNAPHTHA	9316
	PF-WATER	697
	RESIDUE1	74794
	Total product flow	85790
	Reflux ratio	3.48
	Heat duty (Gcal/hr)	
	Condenser duty	-5.22
	Furnace duty	8.21
Pipestil	**Input (Kg/hr)**	
	PP-STM	1260
	STM1	324
	STM2	90
	STM3	72
	RESIDUE1	74794
	Total input	76540
	Product (Kg/hr)	
	HNAPHTHA	9515
	KEROSENE	8344
	DIESEL	2437
	AGO	40758
	RESIDUE2	13749
	PP-WATER	1737
	Total product	76540
	Furnace temperature (°C)	296
	Reflux ratio	5.33
	Heat duty (Gcal/hr)	
	Pumparround1 duty	-1.19
	Pumparround2 duty	-0.17
	Condenser duty)	-6.97
	Furnace duty	6.37

The table above shown that the larger amount of crude was converted into AGO about 47,95%, a value very close to what was predicted initially whie only 1.16 of crude was converted into lights componentes.

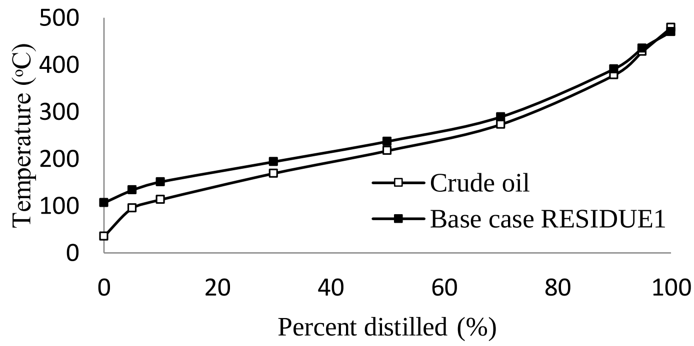


Figure 4.3 TBP curve of blanding oil and base case RESIDUE1

The TBP curves above were made to show the importance of using the pre-fractionator column which removes the light components at relatively low temperatures to avoid cracking as well as to reduce heat energy requirement across furnace and decrease overhead vapor load in the pipestil. From the graphic the RESIDUE1 curve is relatively above the crude oil curve so the initial boiling point was increasing from 35°C to 106°C also the $T_{50\%}$ increased from 217°C to 237°C, which means the biggest quantity of ligth materials was evaporated.

4.2.1 Steady state simulation of the PFD over temperature variation in the preflash furnace

According to Budhiarto (2009) in the crude distillation unit system process variables that affect the crude distillation unit include flash zone temperature of distillation column, top column temperature, top pressure of the distillation column, and the flow rate of stripping steam. The variable in this study is the furnace temperature of preflash column that will affect the flash zone temperature of this column. To perform this simulation, those other variables was maintained constant, it is evident that the furnace temperature changes affect the equilibrium preflash column and the flow rate of product preflash and pipestill, as shown in the following table:

Table 4.3 Result from the simulation over temperature variation in the preflash furnace

Column	Description	Temperature of preflash furnace (°C)				
		260	**285**	**305**	**325**	**345**
Preflash	**Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	Product (Kg/hr)					
	LIGHT	983	840	705	502	476
	LNAPHTHA	9316	10129	10842	11933	12088
	PF-WATER	697	709	717	727	727
	RESIDUE1	74794	74112	73526	72628	72499
	Total product flow	85790	85790	85790	85790	85790
	Reflux ratio	3.48	4.91	5.89	6.53	7.55
	Heat duty (Gcal/hr)					
	Condenser duty	-5.22	-7.24	-8.83	-10.35	-11.84
	Furnace duty	8.21	10.39	12.03	13.54	14.97
Pipestil	**Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	74794	74112	73526	72628	72499
	Total input	76540	75858	75272	74374	74245
	Product (Kg/hr)					
	HNAPHTHA	9515	8640	7912	6829	6655
	KEROSENE	8344	8323	8308	8290	8286
	DIESEL	2437	2971	3351	3836	3942
	AGO	40758	40996	41170	41406	41450
	RESIDUE2	13749	13195	12797	12278	12175
	PP-WATER	1737	1733	1734	1735	1737
	Total product	76540	75858	75272	74374	74245
	Furnace temperature (°C)	296	298	300	302	303
	Reflux ratio	5.33	5.95	6.55	7.64	7.86
	Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.19	-1.18	-1.18	-1.17	-1.17
	Pumparround2 duty	-0.17	-0.17	-0.18	-0.18	-0.18
	Condenser duty)	-6.97	-6.94	-6.91	-6.84	-6.84
	Furnace duty	6.37	6.20	6.13	6.09	6.01

As shown in table 4.3, when the temperature increase from 260°C to 345°C the production of LIGHT ends drops down from 983Kg to 476Kg while the production of LNAPHTHA rose from 9316Kg to 12088Kg. This is due to the fact of keeping consntante the top temperature on preflash column which was 72°C in this simulation. Thus, the larger of the feed temperature that comes to the flash zone of preflash column greater will be the deviation of the temperature between the top and bottom of the column and this increases the amount of condensed liquid. Since the amount of condensate liquid increse so the reflux ratio will increase and the production of LNAPHTHA being higher. Table 4.3 show that increasing temparature cause less production of HNAPHTHA, this is because most naphtha components were removed in the preflash column as LNAPHTHA product this fact was also reported by Handogo, 2011.

4.2.2 Simulation of PFD over variation of strippers position on pipestil column

For this simulation the strippers positions on pipestil column was varied to see its influence on the profile in each of CDU and the product specification. The table 4.4 show the comparison between the simulations over variation on stripper positions when the temperature in the preflash furnace was set at 345°C because is were obtained higher production of DIESEL and KEROSENE the most expensive product and with higher demand see appendix A.3. All those other variable mentioned before was also constant.

Table 4.4 Results from the steady state simulation over variation of strippers positions on pipestil column

Column	**Description**	**Temperature of preflash furnace**			
		345 (°C)			
Preflash	**Input (Kg/hr)**				
	Crude oil	85000	85000	85000	85000
	PF-STM	790	790	790	790
	Total input	85790	85790	85790	85790
	Product (Kg/hr)				
	LIGHT	476	476	476	476
	LNAPHTHA	12088	12088	12088	12088
	PF-WATER	727	727	727	727
	RESIDUE1	72499	72499	72499	72499
	Total product flow	85790	85790	85790	85790

Table 4.4 continued

Column	Description	Temperature of preflash furnace 345 (^oC)			
Preflash	**Reflux ratio**	7.55	7.55	7.55	7.55
	Heat duty (Gcal/hr)				
	Condenser duty	-11.84	-11.84	-11.84	-11.84
	Furnace duty	14.97	14.97	14.97	14.97
Pipestil	**Stripper position**				
	Kerosene stripper	p.6-5	p.5-4	p.6-5	p.5-4
	Diesel stripper	p.14-13	p.14-13	p.13-12	p.13-12
	AGO stripper	p.19-18	p.19-18	p.19-18	p.19-18
	Input (Kg/hr)				
	PP-STM	1260	1260	1260	1260
	STM1	324	324	324	324
	STM2	90	90	90	90
	STM3	72	72	72	72
	RESIDUE1	72499	72499	72499	72499
	Total input	74245	74245	74245	74245
	Product (Kg/hr)				
	HNAPHTHA	6655	6075	6832	6136
	KEROSENE	8286	8175	8309	8186
	DIESEL	3942	7354	6227	8378
	AGO	41450	43911	43731	45264
	RESIDUE2	12175	6992	7408	4542
	PP-WATER	1737	1738	1738	1739
	Total product	74245	74245	74245	74245
	Furnace temperature (^oC)	303	323	315	324
	Reflux ratio	7.86	10.41	8.96	10.41
	Heat duty (Gcal/hr)				
	Pumparround1 duty	-1.17	-1.22	-1.22	-1.26
	Pumparround2 duty	-0.18	-0.21	-0.24	-0.25
	Condenser duty	-6.84	-7.89	-7.89	-8.07
	Furnace duty	6.01	7.25	7.29	7.70

In table 4.4 shows that changing the position of kerosene stripper from p.6-5 to p.5-4 cause a small decrease in the production of KEROSENE, 1.4% while the DIESEL product had a significant increment about 86%. When the diesel stripper positition change from p.14-13 to p.13-12 the kerosene production increase 0.27% and now the increment of DIESEL product was about 58%. Finally changing both kerosene and diesel strippers to p.5-4 and p.13-12 respectively DIESEL product in this case was highest. Shifting up Kerosene Stripper decreases the intersection area of the boiling temperatures between KEROSENE and DIESEL cuts, it mean that those liquid at tray 6 flow to down trays. Thus rise the amount of downward liquid and consequentely flowrate in the diesel side draw and the heat duty in Pipestil furnace increase. Those changes did no affect the quality of products because those side-draw products are taken from trays at which the temperature corresponds to their cut point, also it means that there are dependence between the purity of those cuts and with the composition profile whithin the column (Jobson, 2014).

4.3 Economic analysis

Below are presented the profits obtained from these simulation compared above. Fisrt over the total cost of blending oil per meter cubic was determined using Eq. 21, see table 4.5. Product sales was found by Eq. 22 see table 4.6 and 4.9, it is clear that the case were the strippers positions was at S1p.5-4, S2p.14-13 and S3p.19-18 was highest product sales value about 330.88US$/m^3, see complete comparison profit in appendix A.4. And then, since the LHV of furnace fuel was 47798Kj/Kg, the cost of furnaces heat was determined by Eq. 20 the results are shown in table 4.7 and 4.10. Finally, supply steam cost of 0.43US$/m^3 and preheat cost of 2.4 US$/m^3 was obtained by Eq. 17, see table 4.8 and 4.11. The price of steam is much bigger than the cost of providing cooling water that is why the cost of cooling water was not considered.

Table 4.5 Crude oil purchase cost

Crude oil type	Price($/m^3)	Fraction	Volume(m^3/h)	Cost(US$/h)
Lalang	270.375	0.08	8.288	2240.868
Ramba	271.25	0.3	31.08	8430.45
Duri	263.1875	0.32	33.152	8725.192
Geragai	271.25	0.3	31.08	8430.45
Total	-	1	103.6	27826.96

Oil purchase cost, C_o = 268.6 US$/m^3

Source: Direktorat Jenderal minyak Dan Gas Bumi, 2016

Table 4.6 Product sales for simulation over temperature variariation in the preflash furnace

		Temperature in the preflash furnace(ºC)				
		Base case	**Case2**	**Case3**	**Case4**	**Case5**
Product(Kg/h)	**Price ($/m³)**	**260**	285	305	325	345
LIGHT	285.30	69020	59160	49825	35654	33839
LNAPHTHA	332.58	4673	5072	5418	5947	6020
HNAPHTHA	368.58	4928	4471	4091	3527	3439
KEROSENE	393.23	4683	4672	4664	4656	4652
DIESEL	393.14	1384	1694	1915	2202	2261
AGO	383.23	23220	23354	23458	23592	23615
RESIDUE2	321.59	5930	5692	5518	5293	5248
Total(US$)		113838	104115	94888	80870	79074
Total(US$/m³)		312.91	315.91	319.35	326.25	327.35

Source: Oil Price Information Sistem, 2016

Table 4.7 Cost of furnace heat duty for simulation over temperature variariation in the preflash furnace

	Preflash furnace temperature (ºC)				
	Base case	**Case2**	**Case3**	**Case4**	**Case5**
Description	**260**	**285**	**305**	**325**	**345**
PF heat duty (Kj/hr)	34388739	43516879	50350083	56687819	62677382
PF heat cost (US$/m^3)	2.56	3.23	3.74	4.21	4.66
PPF heat duty (Kj/hr)	26649468	25945621	25677498	25487239	25162668
PPF heat cost (US$/m^3)	1.98	1.93	1.91	1.89	1.87

Source: Oil Price Information Sistem, 2016

Table 4.8 Results from profit comparison for simulation over temperature variariation in the preflash furnace

	Preflash furnace temperature (ºC)				
	Base case	**Case2**	**Case3**	**Case4**	**Optimum case**
Description	**260**	**285**	**305**	**325**	**345**
Product sale (US$/m^3)	312.91	315.91	319.35	326.25	327.35
Cost of oil (US$/m^3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m^3)	0.43	0.43	0.43	0.43	0.43
Cost of heat PF (US$/m^3)	2.56	3.23	3.74	4.21	4.66

Table 4.8 continued

Description	Preflash furnace temperature (ºC)				
	Base case 260	Case2 285	Case3 305	Case4 325	Optimum case 345
Cost of heat PPF (US\$/m^3)	1.98	1.93	1.91	1.89	1.87
Cost of cooling water (US\$/1000m^3)	0.0087	0.0087	0.0087	0.0087	0.0087
Profit (US\$/m^3)	39.33	41.71	44.66	51.11	51.78

In these calculations the annual capital cost was ignored because there's no change in equipment design. After comparing the profit is noted that a highest profit of US\$ 51.78 is found at the furnace preflash temperature of 345ºC (table 4.8). This condiction was considered as optimum case in the study of the influence of temperature on CDU profile.

Table 4.9 Product sales for simulation over variation of strippers position on pipestil column

Product(Kg/h)	Price (\$/m^3)	Temperature in the preflash furnace 345 (ºC)			
		Base case2	Case6	Case7	Case8
		p.6-5 p.14-13 p.19-18	p.5-4 p.14-13 p.19-18	p.6-5 p.13-12 p.19-18	p.5-4 p.13-12 p.19-18
LIGHT	285.30	33839	33839	33839	33839
LNAPHTHA	332.58	6020	6020	6020	6020
HNAPHTHA	368.58	3439	3140	3527	3174
KEROSENE	393.23	4652	4585	4664	4589
DIESEL	393.14	2261	4266	3597	4867
AGO	383.23	23615	24903	24822	25600
RESIDUE2	321.59	5248	3000	3181	1942
Total(US\$)		79074	79753	79650	80031
Total(US\$/m^3)		327.35	330.88	329.19	330.42

Source: Oil Price Information Sistem, 2016

Table 4.10 Cost of furnace heat duty for simulation over variation of strippers position on pipestil column

Description	Preflash furnace temperature 345 (°C)			
	Base case2	**Case6**	**Case7**	**Case8**
Kerosene stripper	p.6-5	p.5-4	p.6-5	p.5-4
Diesel stripper	p.14-13	p.14-13	p.13-12	p.13-12
AGO stripper	p.19-18	p.19-18	p.19-18	p.19-18
PF heat duty (Kj/hr)	62677382	62677382	62677382	62677382
PF heat cost (US$/m^3)	4.66	4.66	4.66	4.66
PPF heat duty (Kj/hr)	25162668	30354300	30708322	32237657
PPF heat cost (US$/m^3)	1.87	2.26	2.28	2.40

Source: Oil Price Information Sistem, 2016

Table 4.11 Results from profit comparison for simulation over variation of strippers position on pipestil column

Description	Preflash furnace temperature 345 (°C)			
	Base case2	**Optimum case2**	**Case7**	**Case8**
Kerosene stripper	p.6-5	p.5-4	p.6-5	p.5-4
Diesel stripper	p.14-13	p.14-13	p.13-12	p.13-12
AGO stripper	p.19-18	p.19-18	p.19-18	p.19-18
Product sale (US$/m^3)	327.35	330.88	329.19	330.42
Cost of oil (US$/m^3)	268.6	268.6	268.6	268.6
Cost of steam (US$/m^3)	0.43	0.43	0.43	0.43
Cost of heat PF (US$/m^3)	4.66	4.66	4.66	4.66
Cost of heat PPF (US$/m^3)	1.87	2.26	2.28	2.40
Cost of cooling water(US$/1000m^3)	0.0087	0.0087	0.0087	0.0087
Profit (US$/m^3)	51.78	54.93	53.22	54.33

The cost of water used was 0.0087 US$/1000m^3 data from BPLDH DKI Jakarta, 2009. Table 4.11 shown that a highest profit of US$ 54.93 is found at the furnace preflash temperature of 345°C, S1p.5-4, S2p.14-13 and S3p.19-18C. It noted also that it wasn't at this conditions of operation that got larger amounts of DIESEL and KEROSENE(table 4.10) the most expensive and higher demand product. This condition was named as optmum case 2 when the infuence of strippers position were analyzed while the first one was called base case 2.

4.4 Dynamic simulation

Since the steady state simulation was done, dynamic simulation was able to be performed. In addition, the pressure units such as valves and pumps, that are not necessary for the steady-state simulation was specified for this simulation. Sizing of the equipment is another requirement of the dynamic simulation. The column diameter, tray spacing, tray active area, weir length, weir height, reflux drum length and diameter are requested for the dynamic simulation of a column. A tray sizing tool can be used to calculate the tray sizes based on flow conditions in the column. After simulating the crude oil distillation process in steady-state by ASPENPlus and entering the parameters required for the dynamic simulation, the files were exported to ASPEN Dynamics. Basic controllers was added after importing the file into ASPEN Dynamics. ASPEN Dynamics provides a number of different types of controllers. The PID Incr. model was used for all controllers in this simulation. The aim was to observe the response of products composition to a change of the feed flow rate, the controlled variable was the temperature of 95% of the ASTM D86 curve representing the composition. The parameters of each controller (gain, integral time and derivative time) were specified.

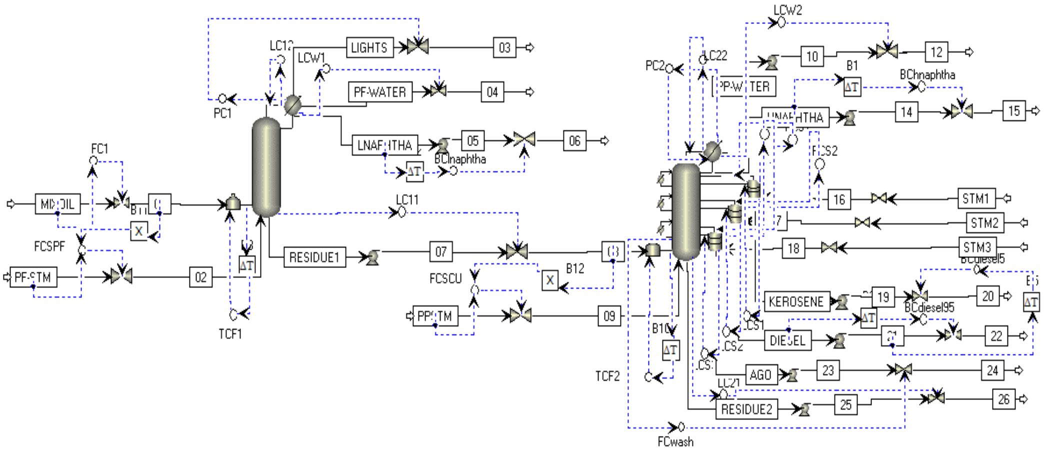


Figure 4.4 PFD from dynamic simulation according to Luyben, 2013

4.5 Analysis of the influence of feed temperature and strippers position on CDU profile

4.5.1 The influence of feed temperature on temperature profile of CDU

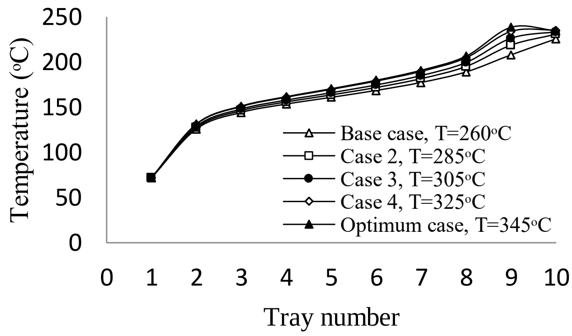
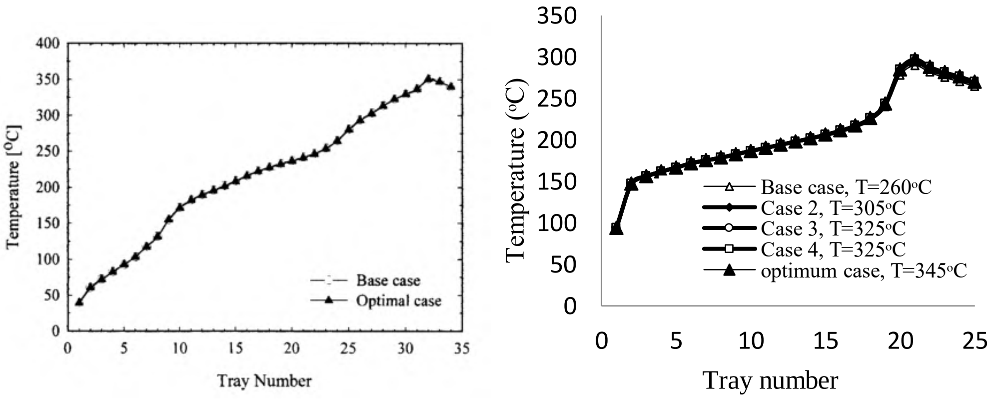


Figure 4.5 Profile of temperature on preflash column



a) Oh et al. (2000) **b) Simulation**

Figure 4.6 Profile of temperature on pipestil column

When the preflash furnace temperature increase from 260°C to 345°C the profile temperature of this column changed substantially starting from tray 3 to 10 and the appearance of a steep in tray 9 (figure 4.5) with a maximum deviation of 38°C between base case curve and the optimum case. Different event occurred in pipestil column were the deviation betweem both case was quity similary and bulk of the temperature changes occurs in a few trays (bottom trays) resulting also in a very steep temperature profile on tray 21 (figure 4.6b). From figures 4.5 and 4.6 it is noted that there is negligible difference between the temperature profiles of the cases when the feeding temperature changed, since the temperature in each stage is a function of the mass flow of the respective stage then it means that in each case the flow it was satisfying and within the limits of tolerances.

The temperature profile of pipestil column was compared with the graph presented by Oh (2000) the results from this comparison is shown in the figures above (4.6a and 4.6b). The results of this validation are satisfactory because is compatible with the literature although there are small discrepancies as the steep localization. This fact is probably linked to four factors: the first is the difference in the number of trays (35 trays for Oh paper and 25 for this research); second difference is the feed tray was 33 and 22 respectivelly; third is the steam and feed temperature and fourth is the stippers' position. The comparison of the temperature profile of preflash column was not possible due to lack of data. As in both columns there was a substantial change in their profile temperature, it mean that there are improvement in fractionation and a consequence changes in product quantity for example the HNAPHTHA for the base case was obtained at 9515kg while with the optimum case was obtained only at 6655kg of HNAPHTHA because increasing feed temperature, the reflux ratio also rise in this case it was from 5.33 to 7.86 (table 4.3) that is why HNAPHTHA product become richer in light components and side draw products thus as RESIDUE2 become rich in heavy components.

4.5.2 The influence of feed temperature on pressure profile of CDU

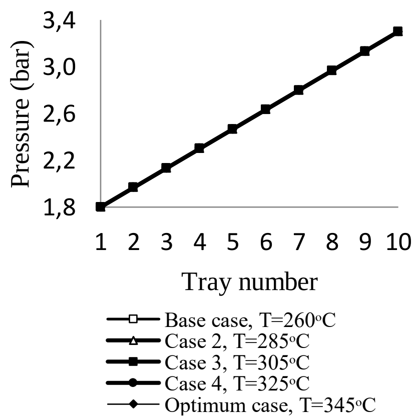


Figure 4.7a Profile of pressure on Preflash column

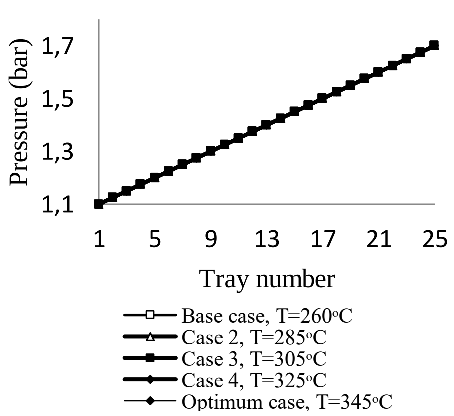


Figure 4.7b Profile of pressure on Pipestil column

The pressure profile of both case in both columns are linear and identical to each other. This means that there are no influences from feed temperature on pressure profile of each columns tray. Since the operation pressure influence grandlly on distillation perfomance, because it have effect on molar flow rates, vapor density, volatility and floding limitations (Liu and Jobson, 1999) for this simulations the preflash and pipestil column was set at 3.3 bar and 1.7 bar respectivelly which worke well for all perfomed simulation, the same range of operation pressure was used also with Luyben, (2013).

4.5.3 The influence of feed temperature on mass and vapor flow profile of CDU

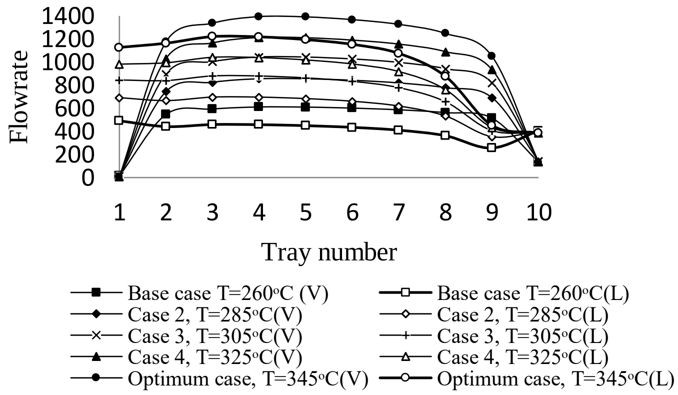
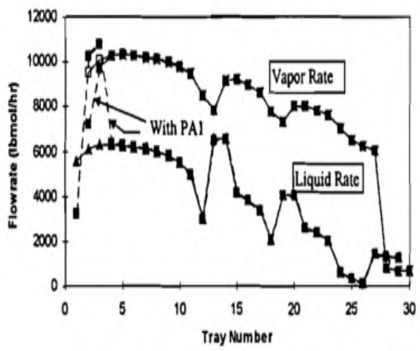
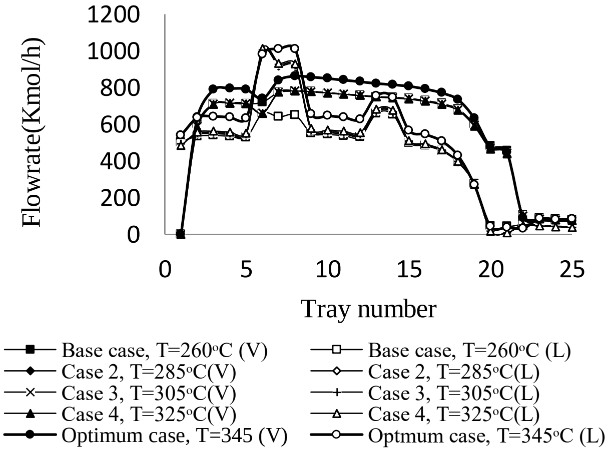


Figure 4.8 Profile of liquid and vapor on each preflash trays





a) Bagajewicz (1997) **b) Simulation**

Figure 4.9 Profile of liquid and vapor on each pipestil trays

The figures above shows the effect of incresing feed temperature on the liquid and vapor tray rates. As shown in Figure 4.8, increasing the feed temperature in the preflash furnace increases the deviation of vapor tray rates between the base case and optimum case due to the high vaporization of feed at preflash column. So, it means that increasing feed temperature, vaporization flow rate and vapor flow rate will increase. The liquid rate of optimum case was higher also than the liquid rate of base case because of the higher reflux ratio, 7.86, since the vapor ascend must be equal to the descendent liquid to ensure mass exchange between them, so the reflux must be higher to maintain the stability of temperature, pressure and internal reflux in the column. The same behaviour was verified on pipestil column but the liquid rate line for both case has a steep in tray 6 and 13 because is at this tray where the pumparround liquid return to the main column. For both columns there is no a disturbance in its operation caused by feed temperature changes, these deviation is normal and the column operation was satisfying. The similar mass flow profile of pipestil column was presented by Bagajewicz (1997) as shown in the figure 4.9a. There are small discrepancies between this liquid and vapor profile due to the differences in location of pumparounds and number o column trays. Bagajewicz research the liquid draw from tray 15 through pumparound2 and it return to the pipestil column on tray 13, the liquid also is taken from tray 21 and it back to the main column on tray 19 while in this research the liquid draw from tray 8 through pumparound1 and it back on tray 6, and then the liquid is taken from tray 14 through pumparound2 and it back on tray 13 that is why there is diference in steep localization.

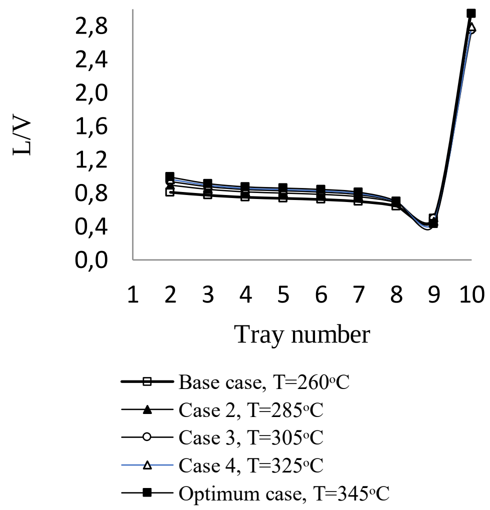


Figure 4.10a Ratio between liquid living and vapor arriving in each preflash trays

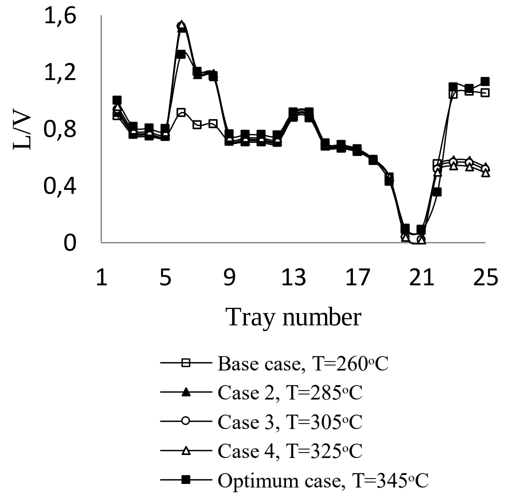
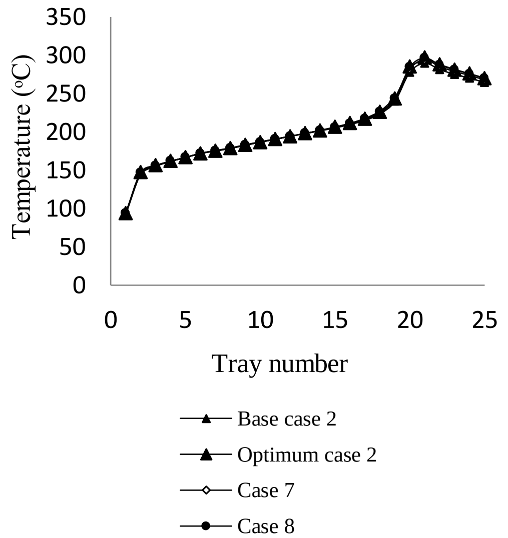


Figure 4.10b Ratio between liquid living and vapor arriving in each pipestil trays

The Figures 4.10a and 4.10b clearly shows the effect of increasing temperature on the ratio of liquid and vapor (L/V). Notice that the liquid vapor ratio of optimum case in the preflash column become higher from tray 8 until the top column, while in pipestil column it become higher from tray 19. It means that fractionation was improved, however, volume of liquid which descends along the column is greater thus thereby ensure improvement in internal reflux and minimized the overvaporized materials without compromising the quality of product. As stated by Silvestre (2005), this volume of liquid will be larger as the temperature of the feed increases and if the temperature increase extremely, it may increase the pressure drop between the trays causing floding and cracking of light materials in furnace.

4.5.4 The influence of strippers' position on CDU profile



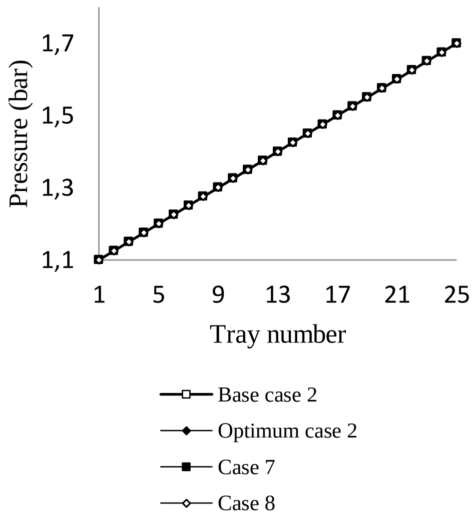
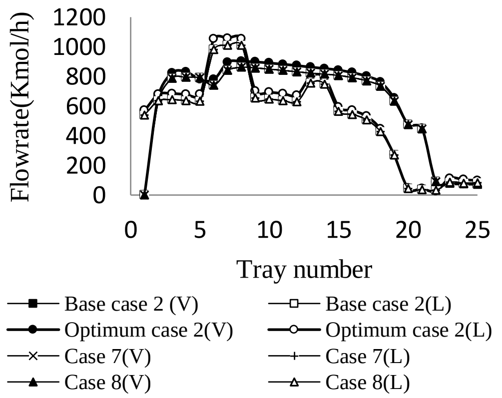


Figure 4.11a Temperature profile of pipestil pipestil column

Figure 4.11b Pressure profile of pipestil column



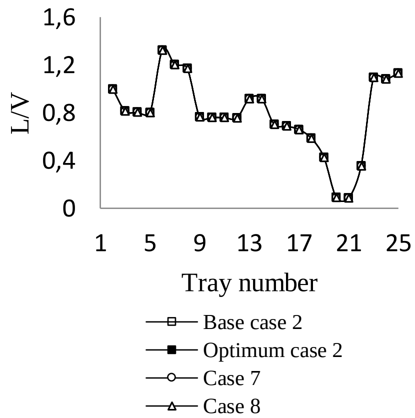


Figure 4.11c Profile of liquid and vapor flow on each stage of pipestil column

Figure 4.11d Ratio between liquid living and vapor arriving in each tray of pipestil column

Displacing kerosene stripper from p.6-5 to p.5-4 has no effect o pressure profile (see figure 4.11b), but it allows the liquid from tray 6 to flow normally down to trays below, thus incrising the rate of downflow liquid in overal column as shown in figure 4.11c. This increment of downflow liquid cause an increament of heat duty in the pipestil furnace because the need of higher temperature to ensure a vaporization of back liquid, that is why the temperature profile on the botton trays was high see figure 4.11b. All these have contributed to bulk temperature changes to be higher in the pipestil column as shown in figure 4.11a. There is no significant changes on CDU profile when strippers position was changed and it not effect the CDU operation.

Table 4.12 Base case 2 side draw product specification

Variable	Kerosene	Diesel	AGO
Stripper position	p.6-5	p.14-13	p.19-18
Temperature (^oC)	132.3	170.9	237.6
Pressura (bar)	1.2	1.4	1.5
Enthalpy (Gcal/hr)	-3.2	-1.4	-13.1
Specific gravity	0.799	0.813	0.847
API gravity	45.7	42.6	35.5
95% vol. bp. ASTM D86	193	222	246.3

Table 4.13 Optimum case 2 side draw product specification

Variable	Kerosene	Diesel	AGO
Stripper position	p.5-4	p.14-13	p.19-18
Temperature (^oC)	128.9	180.3	247
Pressura (bar)	1.2	1.4	1.5
Enthalpy (Gcal/hr)	-3.1	-2.6	-13.8
Specific gravity	0.797	0.814	0.855
API gravity	46.1	42.4	33.9
95% vol. bp. ASTM D86	191.7	222	277

Tabel 4.12 and 4.13 shown the side draw product specification of both cases. The comparison of product clearly shown that small deviaviation in temperature, density occur between them, onle AGO had a highest deviation about 30,7 ^oC in the boiling point. Those changes did no affect the quality of kerosene,

diesel and AGO because those side-draw products are taken from trays at which the temperature corresponds to their cut point, accordingg to Leffler (2008) kerosene cut range is $157^{o}C$ - $232\,^{o}C$, disel cut range $232\,^{o}C$ - $343\,^{o}C$ and AGO cut range $343\,^{o}C$ - $454\,^{o}C$.

4.6 Dynamic simulation result

The products mass flow was observed after increasing and reducing the feed flow of crude oil by 20%. Mass flow of the products was changed until the composition (ASTM D86 95 % boiling point) reached the value corresponding to the given product as shown in figures below:

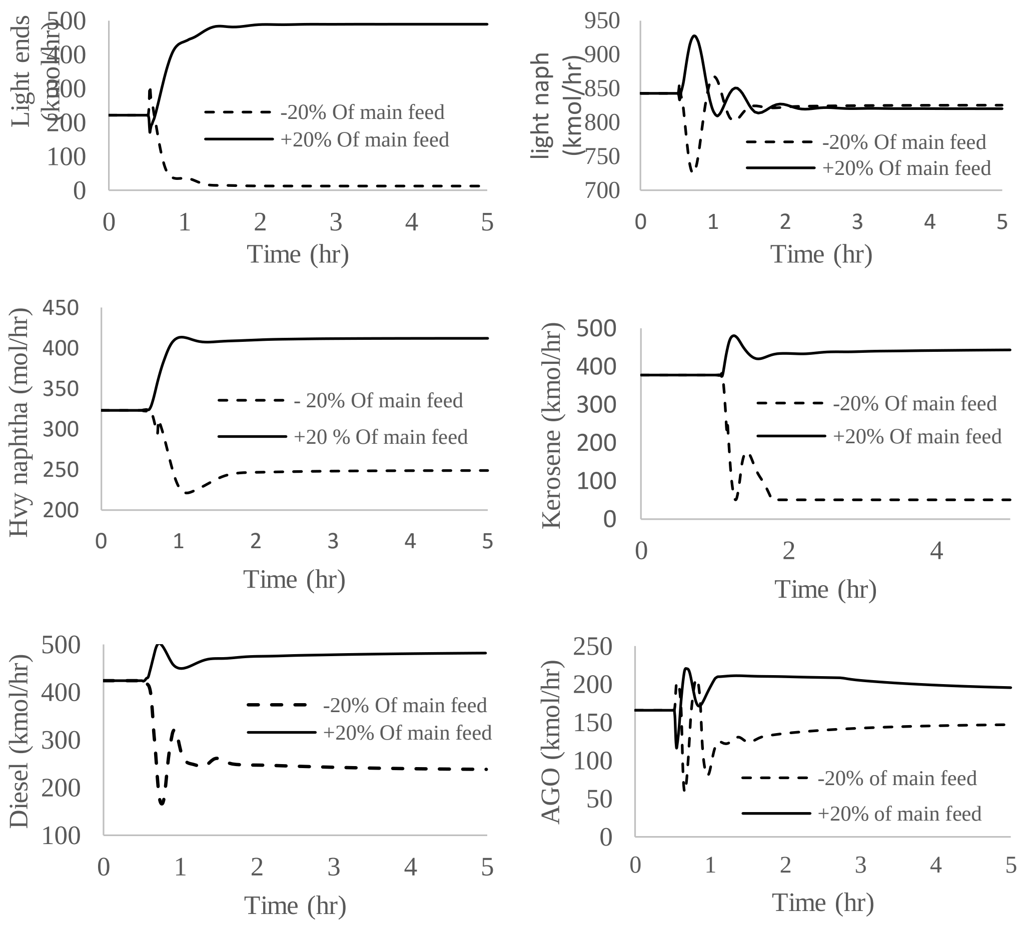
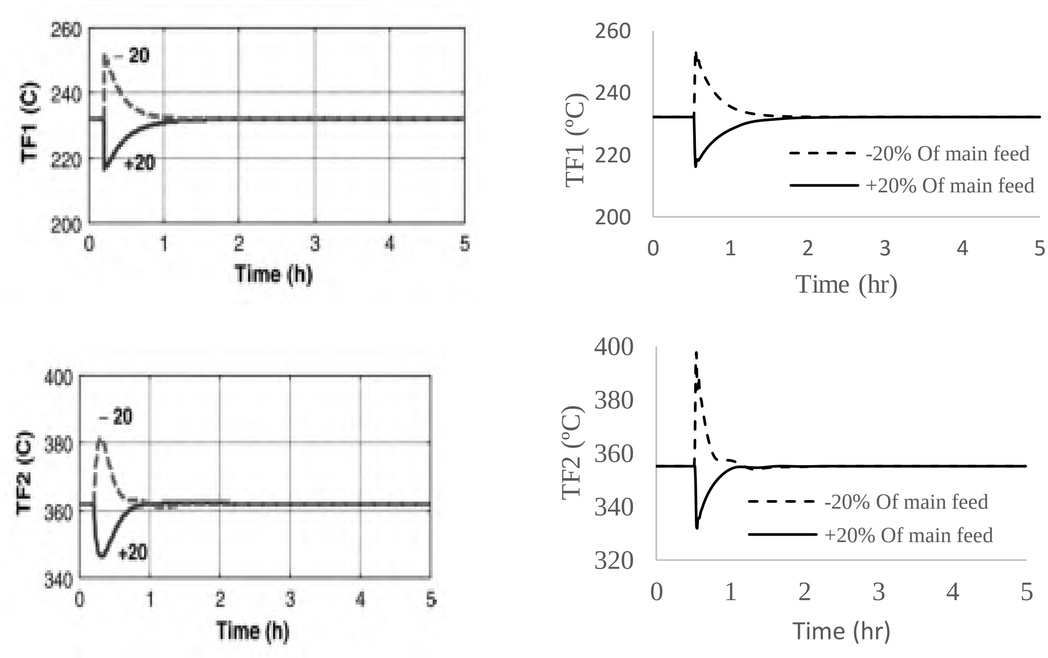


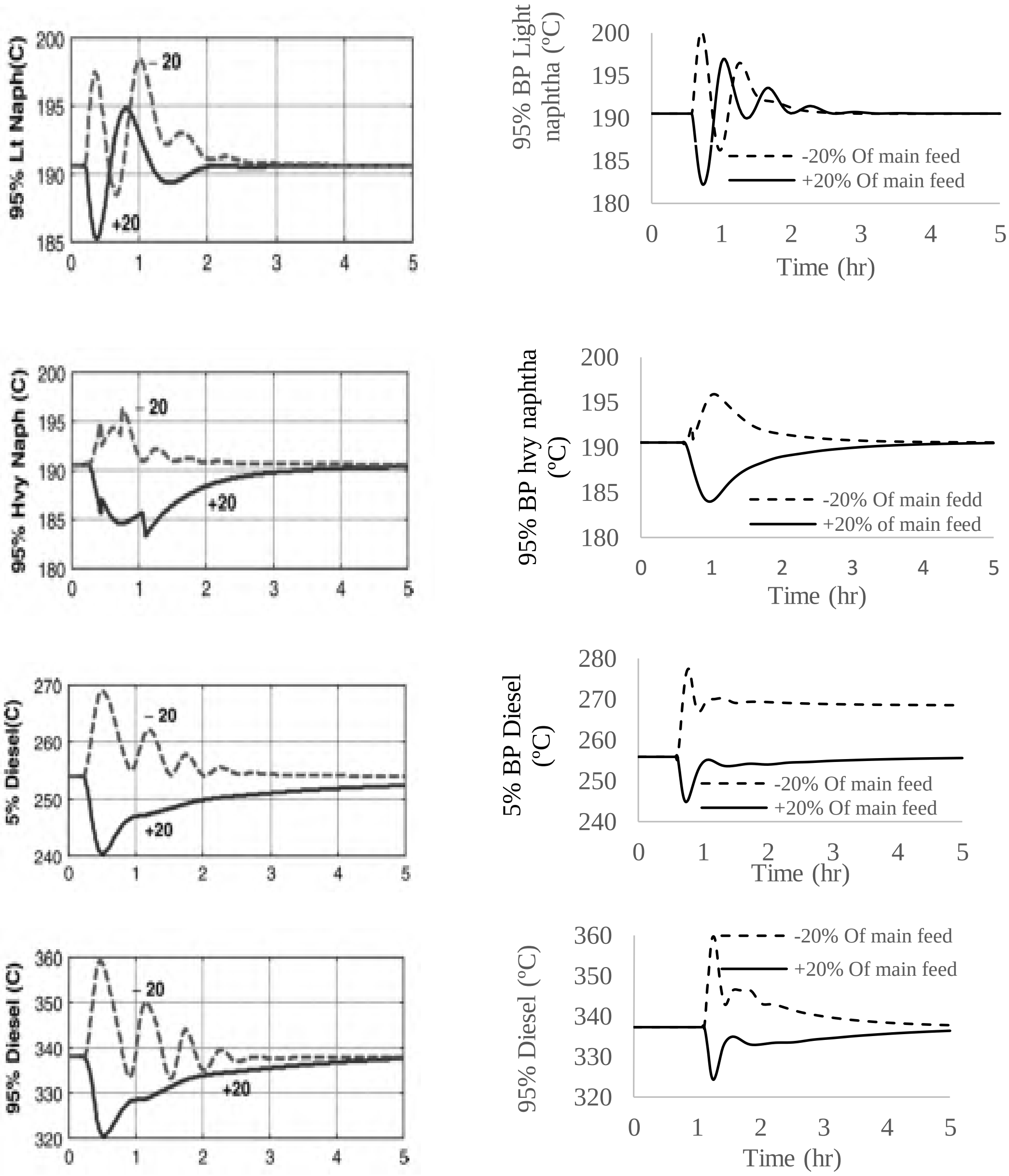
Figure 4.12 Behavior of CDU products flow

Figures above shows the behavior of the CDU the disturbance of both positive and negative 20% change are given in the setpoint of the main feed flow control. Increasing the mass flow of the feed by 20%, the mass flow of the lights ends (figure 4.12a) and heavy naphtha (figure 4.12c) reached a new steady-state at higher and low rate flows; however, the mass flow of light naphtha and diesel flow rate unexpectedly, although slightly, decreased as shown in figure 4.12b and 4.12d respectively. The time necessary to reach a new steady-state for all product flow rate was at 1.5 hr, the same event was reported by Hardary and Pavlik (2009). The deviation of this method was significantly lower than that of the second method. Experimental data for the simulation results verification in dynamic mode were not available. Some limitation of ASPEN Dynamics in supporting different types of column configurations and in dynamic simulation of the process start were found (Aspen Plus, 2006).



a) Luyben (2013) **b) Simulation**

Figure 4.13 Temperature controllers



a) Luyben (2013) **b) Simulation**

Figure 4.13 Temperature controllers, continued

The naphtha (light and heavy) and diesel flow was manipulated to control the 95% boiling point at 190ºC and 222ºC respectively also the 5% boiling point of diesel was controlled but now manipulating kerosene flow (Figure 4.4). The disturbance of both positive and negative 20% change are given in the setpoint of the main feed flow controller at time equal to 1hr, the changes are shown in figure 4.13. The figures above shown that the deviations in the 5% and 95% boiling points of diesel product are about 20 °C for these disturbances relatively large while the maximum deviation in the 95% boiling point of naphtha products are about 5%. The result are in agreement with these result reported by Luyben (2013) as illustrated in figures 4.13.

Table 4.14 Integral of the time –weighted absolute error (ITAE), based on feed disturbance

	Base case 2		**Optimum case 2**	
Controller	**-20%**	**20%**	**-20%**	**20%**
TC furnace 1	427	461	427	461
TC furnace 2	318	445	318	445
95% BP ASTM D86 L-naphtha	834	735	834	732
95% BP ASTM D86 H-naphtha	847	1421	941	1302
5% BP ASTM D86 Diesel	15826	133473	17100	1461
95% BP ASTM D86 Diesel	3248	3003	3367	2884

The table 4.14 shown the tuning relation based on the time-weighted absolute error (ITAE), where the error signal *e(t)* is the difference between the set point and the measurement. Then the ITAE is calculated using the equation:

$$\mathrm{ITAE} = \int_0^\infty |e(t)dt$$

In this equation the time *(t)* is used because the initial error for step response in this research is large so was reasonable to weight this error in both base case 2 and optimum case 2 to see the case that give a minimum ITAE. The table 4.14 shown that the ITAE in the 95% boiling point of light naphtha and the TCF1 are similar in both cases may because this controllers are linked on preflash column since no change in configuration was made in this column. While the ITAE in the 95% boiling point of diesel have a deviation about have a small deviation about 100, it does not affect da quality of diesel.

APPENDIX

A.1 True boiling point (TBP) curve of product yields

Column characteristic	Preflash column	Pipestil column
Section starting stage (m)	2	2
Section ending stage (m)	10	25
Column diameter (m)	3.42	3.36
Column height (m)	5.86	16.8
Side downcomer velocity (m/s)	0.065	0.071
Side weir length (m)	2.49	2.44
Side downcomer width (m)	0.54	0.53
Flow path length (m)	2.35	2.31

A.2 True boiling point (TBP) curve of product yields

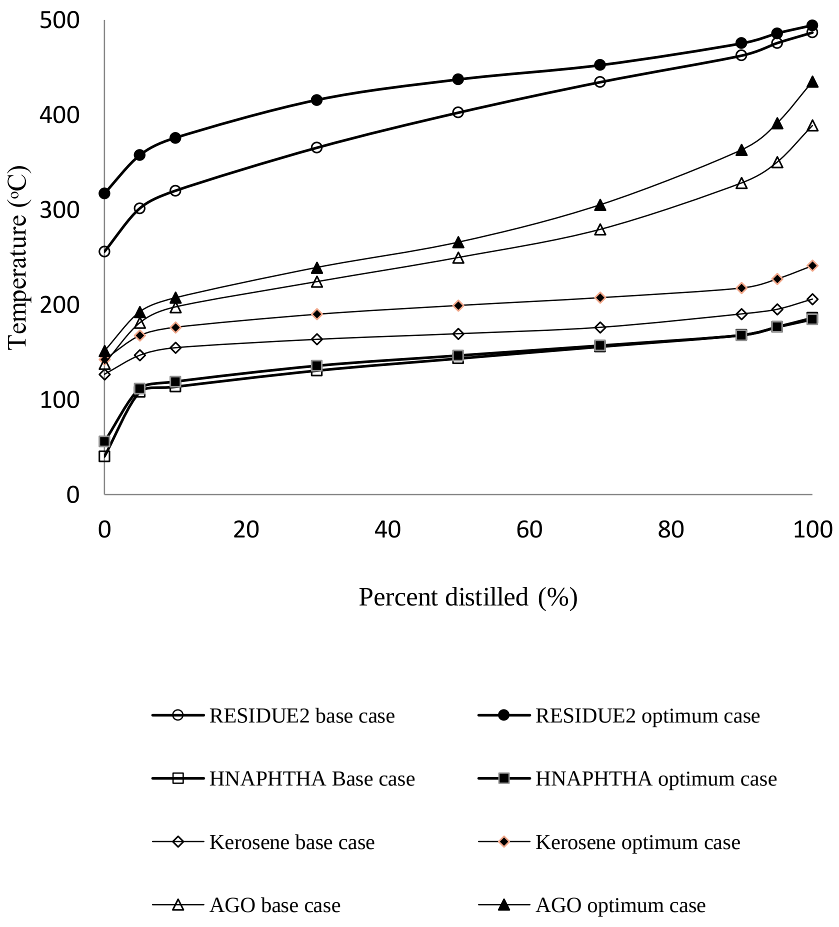


Figure A.2 TBP curve of product yields

A.3 cut range boiling point of base case and optimum case

Table A.3a cut range boiling point of base case

Product yield	Cut boiling point range ASTM D86 (°C)
LIGHTS	3.3 – 120.2
LNAPHTHA	28.9 – 130.8
HNAPHTHA	50.3 – 153.1
KEROSENE	125.4 – 177.1
DIESEL	147.5 – 206.8
ATG	164.8 – 356.9
RESIDUE2	292.3 – 458.9

Table A.3b cut range boiling point of optimum case

Product yield	Cut boiling point range ASTM D86 (°C)
LIGHTS	3.3 – 120.2
LNAPHTHA	27.5 – 129.4
HNAPHTHA	52.7 – 151.5
KEROSENE	124.1 – 176.4
DIESEL	145.2 – 206.6
ATG	164.8 – 356.9
RESIDUE2	302.5 – 458.8

A.3 Results from the steady state simulation with variation strippers positions on pipestil column

Table A.3 Results from simulation with variation strippers positions on pipestil

Column	Description	Temperature of preflash furnace (°C)				
		Base case	**345**	**345**	**345**	**345**
Stripper	Kerosene stripper	p.6-5	p.6-5	p.6-5	p.6-5	p.6-5
	Diesel stripper	p.14-13	p.14-13	p.13-12	p.12-11	p.14-13
	AGO stripper	p.19-18	p.19-18	p.19-18	p.19-18	p.20-19
Preflash	**a) Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	b) Product (Kg/hr)					
	LIGHT	983	476	476	476	476
	LNAPHTHA	9316	12088	12088	12088	12088
	PF-WATER	697	727	727	727	727
	RESIDUE1	74794	72499	72499	72499	72499
	Total product	85790	85790	85790	85790	85790
	c) Reflux ratio	3.48	7.55	7.55	7.55	7.55
	d) Heat duty (Gcal/hr)					
	Condenser duty	-5.22	-11.84	-11.84	-11.84	-11.84
	Furnace duty	8.21	14.97	14.97	14.97	14.97
Pipestil	**a) Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	74794	72499	72499	72499	72499
	Total input	76540	74245	74245	74245	74245
	b) Product (Kg/hr)					
	HNAPHTHA	9515	6655	6832	6957	6816
	KEROSENE	8344	8286	8309	8341	8294
	DIESEL	2437	3942	6227	7788	6115
	AGO	40758	41451	43731	45835	43622
	RESIDUE2	13749	12175	7408	3586	7661
	PP-WATER	1738	1737	1738	1738	1738
	Total product	76540	74245	74245	74245	74245
	c) Furnace temperature (°C)	296	303	315	339	319
	d) Reflux ratio	5.33	7.86	8.96	10.04	8.88
	e) Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.19	-11.17	-1.22	-1.27	-1.20
	Pumparround2 duty	10.17	-0.18	-0.24	-0.30	-0.20
	Condenser duty	-6.97	-6.84	-7.89	-8.68	-7.70
	Furnace duty	6.37	6.01	7.29	8.42	7.22

Table A.3 continued

Column	Description	Temperature of preflash furnace (^oC)				
		345	345	**345**	**345**	345
Stripper	Kerosene stripper	p.6-5	p.6-5	p.6-5	p.6-5	p.6-5
	Diesel stripper	p.15-14	p.14-13	p.13-12	p.13-12	p.12-11
	AGO stripper	p.20-19	p.18-17	p.20-19	p.18-17	p.18-17
Preflash	**c) Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	d) Product (Kg/hr)					
	LIGHT	476	476	476	476	476
	LNAPHTHA	12088	12088	12088	12088	12088
	PF-WATER	727	727	727	727	727
	RESIDUE1	72499	72499	72499	72499	72499
	Total product	85790	85790	85790	85790	85790
	c) Reflux ratio	7.55	7.55	7.55	7.55	7.55
	d) Heat duty (Gcal/hr)					
	Condenser duty	-11.84	-11.84	-11.84	-11.84	-11.84
	Furnace duty	14.97	14.97	14.97	14.97	14.97
Pipestil	**c) Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	72499	72499	72499	72499	72499
	Total input	74245	74245	74245	74245	74245
	d) Product (Kg/hr)					
	HNAPHTHA	6761	6497	6928	6758	6919
	KEROSENE	8279	8278	8314	8304	8338
	DIESEL	5532	1656	7458	5205	7277
	AGO	42923	39694	45339	42582	45067
	RESIDUE2	9012	16384	4467	9658	4906
	PP-WATER	1738	1736	1738	1737	1738
	Total product	74245	74245	74245	74245	74245
	c) Furnace temperature (^oC)	314	290	334	312	333
	d) Reflux ratio	8.59	7.06	9.75	8.41	9.66
	e) Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.18	-1.14	-1.24	-1.20	-1.26
	Pumparround2 duty	-0.17	-0.16	-0.25	-0.23	-0.30
	Condenser duty	-7.45	-6.17	-8.44	-7.32	-8.36
	Furnace duty	6.87	5.25	8.12	6.76	8.05

Table A.3 continued

Column	Description	Temperature of preflash furnace (^oC)				
		345	**345**	**345**	**345**	**345**
Stripper	Kerosene stripper	p.5-4	p.5-4	p.5-4	p.5-4	p.5-4
	Diesel stripper	p.15-14	p.14-13	p.14-13	p.14-13	p.13-12
	AGO stripper	p.20-19	p.19-18	p.20-19	p.18-17	p.19-18
Preflash	**e) Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	f) Product (Kg/hr)					
	LIGHT	476	476	476	476	476
	LNAPHTHA	12088	12088	12088	12088	12088
	PF-WATER	727	727	727	727	727
	RESIDUE1	72499	72499	72499	72499	72499
	Total product	85790	85790	85790	85790	85790
	c) Reflux ratio	7.55	7.55	7.55	7.55	7.55
	d) Heat duty (Gcal/hr)					
	Condenser duty	-11.84	-11.84	-11.84	-11.84	-11.84
	Furnace duty	14.97	14.97	14.97	14.97	14.97
Pipestil	**e) Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	72499	72499	72499	72499	72499
	Total input	74245	74245	74245	74245	74245
	f) Product (Kg/hr)					
	HNAPHTHA	6062	6075	6175	5826	6136
	KEROSENE	8166	8175	8176	8169	8186
	DIESEL	7174	7354	8700	3640	8378
	AGO	43722	43911	45731	40444	45263
	RESIDUE2	7382	6992	3724	14428	4542
	PP-WATER	1739	1739	1739	1737	1739
	Total product	74245	74245	74245	74245	74245
	c) Furnace temperature (^oC)	321	323	338	295	324
	d) Reflux ratio	10.27	1041	11.43	8.49	10.41
	e) Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.21	-1.22	-1.24	-1.17	-1.26
	Pumparround2 duty	-0.17	-0.21	-0.22	-0.16	-0.25
	Condenser duty	-7.79	-7.89	-8.66	-6.46	-8.07
	Furnace duty	7.30	7.25	8.34	5.65	7.70

Table A.3 continued

Column	Description	Temperature of preflash furnace (^oC)				
		345	**345**	**345**	**345**	**345**
Stripper	Kerosene stripper	p.5-4	p.5-4	p.5-4	p.5-4	p.5-4
	Diesel stripper	p.13-12	p.13-12	p.12-11	p.12-11	p.15-14
	AGO stripper	p.20-19	p.18-17	p.19-18	p.18-17	p.19-18
Preflash	**g) Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	h) Product (Kg/hr)					
	LIGHT	476	476	476	476	476
	LNAPHTHA	12088	12088	12088	12088	12088
	PF-WATER	727	727	727	727	727
	RESIDUE1	72499	72499	72499	72499	72499
	Total product	85790	85790	85790	85790	85790
	c) Reflux ratio	7.55	7.55	7.55	7.55	7.55
	d) Heat duty (Gcal/hr)					
	Condenser duty	-11.84	-11.84	-11.84	-11.84	-11.84
	Furnace duty	14.97	14.97	14.97	14.97	14.97
Pipestil	**g) Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	72499	72499	72499	72499	72499
	Total input	74245	74245	74245	74245	74245
	h) Product (Kg/hr)					
	HNAPHTHA	6174	5826	6171	6131	5806
	KEROSENE	8188	8169	8204	8202	8162
	DIESEL	8951	3640	9204	8615	3398
	AGO	46132	40444	46502	45585	40278
	RESIDUE2	3061	14428	2425	3972	14863
	PP-WATER	1739	1737	1739	1739	1737
	Total product	74245	74245	74245	74245	74245
	c) Furnace temperature (^oC)	341	295	345	338	294
	d) Reflux ratio	11.63	8.49	11.90	11.38	8.43
	e) Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.27	-1.17	-1.30	-1.29	-1.16
	Pumparround2 duty	-0.26	-0.16	-0.30	-0.31	-0.13
	Condenser duty	-8.79	-6.46	-8.95	-8.57	-6.40
	Furnace duty	8.53	5.65	8.76	8.32	5.55

Table A.3 continued

Column	Description	Temperature of preflash furnace (^oC)				
		345	**345**	**345**	**345**	**345**
Stripper	Kerosene stripper	p.4-3	p.4-3	p.4-3	p.4-3	p.4-3
	Diesel stripper	p.15-14	p.14-13	p.14-13	p.13-12	p.13-12
	AGO stripper	p.20-19	p.19-18	p.18-17	p.19-18	p.18-17
Preflash	**i) Input (Kg/hr)**					
	Crude oil	85000	85000	85000	85000	85000
	PF-STM	790	790	790	790	790
	Total input	85790	85790	85790	85790	85790
	j) Product (Kg/hr)					
	LIGHT	476	476	476	476	476
	LNAPHTHA	12088	12088	12088	12088	12088
	PF-WATER	727	727	727	727	727
	RESIDUE1	72499	72499	72499	72499	72499
	Total product	85790	85790	85790	85790	85790
	c) Reflux ratio	7.55	7.55	7.55	7.55	7.55
	d) Heat duty (Gcal/hr)					
	Condenser duty	-11.84	-11.84	-11.84	-11.84	-11.84
	Furnace duty	14.97	14.97	14.97	14.97	14.97
Pipestil	**i) Input (Kg/hr)**					
	PP-STM	1260	1260	1260	1260	1260
	STM1	324	324	324	324	324
	STM2	90	90	90	90	90
	STM3	72	72	72	72	72
	RESIDUE1	72499	72499	72499	72499	72499
	Total input	74245	74245	74245	74245	74245
	j) Product (Kg/hr)					
	HNAPHTHA	5296	5283	5040	5300	1102
	KEROSENE	8040	8043	8041	8048	1729
	DIESEL	8947	9120	5597	9947	1908
	AGO	44678	44860	41159	46030	9833
	RESIDUE2	5545	5198	12670	3179	1332
	PP-WATER	1740	1740	1739	1740	16177
	Total product	74245	74245	74245	74245	74245
	c) Furnace temperature (^oC)	329	332	301	341	325
	d) Reflux ratio	12.73	12.95	10.57	13.77	12.48
	e) Heat duty (Gcal/hr)					
	Pumparround1 duty	-1.21	-1.23	-1.18	-1.26	-1.24
	Pumparround2 duty	-0.16	-0.20	-0.16	-0.25	-0.24
	Condenser duty	-8.24	-8.34	-6.76	-8.81	-7.97
	Furnace duty	7.81	7.95	6.04	8.53	7.57

Table A.3 continued

Column	Description	Temperature of preflash furnace (^oC)	
		345	**345**
Stripper	Kerosene stripper	p.4-3	p.4-3
	Diesel stripper	p.12-11	p.15-14
	AGO stripper	p.18-17	p.19-18
Preflash	**k) Input (Kg/hr)**		
	Crude oil	85000	85000
	PF-STM	790	790
	Total input	85790	85790
	l) Product (Kg/hr)		
	LIGHT	476	476
	LNAPHTHA	12088	12088
	PF-WATER	727	727
	RESIDUE1	72499	72499
	Total product	85790	85790
	c) Reflux ratio	7.55	7.55
	d) Heat duty (Gcal/hr)		
	Condenser duty	-11.84	-11.84
	Furnace duty	14.97	14.97
Pipestil	**k) Input (Kg/hr)**		
	PP-STM	1260	1260
	STM1	324	324
	STM2	90	90
	STM3	72	72
	RESIDUE1	72499	72499
	Total input	74245	74245
	l) Product (Kg/hr)		
	HNAPHTHA	5257	5045
	KEROSENE	8055	8038
	DIESEL	10193	5404
	AGO	46327	41029
	RESIDUE2	2672	12991
	PP-WATER	1740	1739
	Total product		
	c) Furnace temperature (^oC)	344	300
	d) Reflux ratio	14.07	10.48
	e) Heat duty (Gcal/hr)		
	Pumparround1 duty	-1.29	-1.17
	Pumparround2 duty	-0.31	-0.12
	Condenser duty	-8.90	-6.72
	Furnace duty	8.69	5.95

A.4 Results from profit comparison

Table A.4 calculated profit

Description	Preflash furnace temperature (^oC)				
	Base case	**345**	**345**	**345**	**345**
Kerosene stripper	p.6-5	p.6-5	p.6-5	p.6-5	p.6-5
Diesel stripper	p.14-13	p.14-13	p.13-12	p.12-11	p.14-13
AGO stripper	p.19-18	p.19-18	p.19-18	p.19-18	p.20-19
Product sale (US$/m3)	312.91	327.35	329.19	330.65	329
Cost of oil (US$/m3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m3)	0.43	0.43	0.43	0.43	0.43
Heat Cost of preflash furnace (US$/m3)	2.56	4.66	4.66	4.66	4.66
Heat Cost of preflash furnace (US$/m3)	1.98	1.87	2.28	2.62	2.25
Profit (US$/m3)	39.33	51.79	53.23	54.34	53.19

Table A.4 continued

Description	Preflash furnace temperature (^oC)				
	345	**345**	**345**	**345**	**345**
Kerosene stripper	p.6-5	p.6-5	p.6-5	p.6-5	p.6-5
Diesel stripper	p.14-13	p.13-12	p.13-12	p.12-11	p.15-14
AGO stripper	p.18-17	p.20-19	p.18-17	p.18-17	p.20-19
Product sale (US$/m3)	326	330	328	330	329
Cost of oil (US$/m3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m3)	0.43	0.43	0.43	0.43	0.43
Heat Cost of preflash furnace (US$/m3)	4.66	4.66	4.66	4.66	4.66
Heat Cost of preflash furnace (US$/m3)	1.63	2.53	2.10	2.50	2.14
Profit (US$/m3)	50.34	54.12	52.52	53.94	52.77

Table A.4 continued

Description	Preflash furnace temperature (^oC)				
	345	**345**	**345**	**345**	**345**
Kerosene stripper	p.5-4	p.5-4	p.5-4	p.5-4	p.5-4
Diesel stripper	p.14-13	p.14-13	p.14-13	p.13-12	p.13-12
AGO stripper	p.19-18	p.20-19	p.18-17	p.19-18	p.20-19
Product sale (US$/m3)	330.88	331	327	330.42	331
Cost of oil (US$/m3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m3)	0.43	0.43	0.43	0.43	0.43
Heat Cost of preflash furnace (US$/m3)	4.66	4.66	4.66	4.66	4.66
Heat Cost of preflash furnace (US$/m3)	2.26	2.59	1.76	2.40	2.66
Profit (US$/m3)	54.93	54.46	51.11	54.34	54.65

Table A.4 continued

Description	Preflash furnace temperature (°C)				
	345	**345**	**345**	**345**	**345**
Kerosene stripper	p.5-4	p.5-4	p.5-4	p.5-4	p.5-4
Diesel stripper	p.13-12	p.12-11	p.12-11	p.15-14	p.15-14
AGO stripper	p.18-17	p.19-18	p.18-17	p.19-18	p.20-19
Product sale (US$/m3)	327	331.22	331	326	329
Cost of oil (US$/m3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m3)	0.43	0.43	0.43	0.43	0.43
Heat Cost of preflash furnace (US$/m3)	4.66	4.66	4.66	4.66	4.66
Heat Cost of preflash furnace (US$/m3)	1.76	2.73	2.59	1.73	2.27
Profit (US$/m3)	51.11	54.80	54.34	50.98	53.40

Table A.4 continued

Description	Preflash furnace temperature (°C)				
	345	**345**	**345**	**345**	**345**
Kerosene stripper	p.4-3	p.4-3	p.4-3	p.4-3	p.4-3
Diesel stripper	p.14-13	p.14-13	p.13-12	p.13-12	p.12-11
AGO stripper	p.19-18	p.18-17	p.19-18	p.18-17	p.18-17
Product sale (US$/m3)	330.31	327	331.08	330	331
Cost of oil (US$/m3)	268.6	268.6	268.6	268.6	268.6
Cost of steam (US$/m3)	0.43	0.43	0.43	0.43	0.43
Heat Cost of preflash furnace (US$/m3)	4.66	4.66	4.66	4.66	4.66
Heat Cost of preflash furnace (US$/m3)	2.47	1.88	2.47	2.35	2.71
Profit (US$/m3)	54.15	51.83	54.91	53.72	54.86

Table A.4 continued

Description	Preflash furnace temperature (°C)				
	345	**345**			
Kerosene stripper	p.4-3	p.4-3			
Diesel stripper	p.15-14	p.15-14			
AGO stripper	p.19-18	p.20-19			
Product sale (US$/m3)	327	330			
Cost of oil (US$/m3)	268.6	268.6			
Cost of steam (US$/m3)	0.43	0.43			
Heat Cost of preflash furnace (US$/m3)	4.66	4.66			
Heat Cost of preflash furnace (US$/m3)	1.85	2.43			
Profit (US$/m3)	51.74	54.08			

A.5 Mass and vapor flow on each column stages

Table A.5c Mass and vapor flow on each stage of pipestil column (Base case)

Stage	Liquid flow on stage (kmol/hr):	Vapor flow on stage (kmol/hr):	Mass flow liquid from stage (kg/hr):	Mass flow vapor to stage (kg/hr):	Volume flow liquid from stage (m^3/hr):	Volume flow vapor to stage (m^3/hr):
1	503	0	61980	61980	86,5	17926
2	535	600	67781	79033	101	21256
3	539	711	70422	81674	106	21217
4	535	715	71492	82744	107	20890
5	529	711	72042	83295	109	20940
6	659	721	136800	95383	207	22213
7	640	774	126588	97259	192	22103
8	650	780	127419	98091	193	21764
9	552	777	79044	98316	120	21329
10	545	770	79174	98447	120	20899
11	538	763	79240	98513	121	20479
12	531	756	79236	98509	121	20065
13	659	749	99532	100734	152	20451
14	653	746	99829	99803	153	19412
15	500	740	77702	99321	119	18959
16	484	729	76568	98188	117	18419
17	454	713	73944	95563	113	17673
18	393	683	67229	88849	102	16660
19	275	600	52712	71399	79,4	13507
20	47,0	482	4777	67082	7,04	12814
21	42,0	454	2373	64679	3,47	12626
22	57,1	104	17105	4616	23,9	2320
23	88,3	84,8	15960	3470	21,9	2118
24	84,1	79,2	15162	2673	20,6	1973
25	79,5	75,6	13749	1260	18,4	1941

Table A.5d Mass and vapor flow on each stage of pipestil column (Optimum case)

Stage	Liquid flow on stage (kmol/hr):	Vapor flow on stage (kmol/hr):	Mass flow liquid from stage (kg/hr):	Mass flow vapor to stage (kg/hr):	Volume flow liquid from stage (m^3/hr):	Volume flow vapor to stage (m^3/hr):
1	570	0	71056	71056	99,0	20054
2	678	667	86380	94194	129,1	24747
3	684	825	89480	97294	134,5	24714
4	679	831	90731	98546	136,7	24859
5	678	780	92226	95936	139,4	22713
6	1052	779	145642	112705	220,5	25760
7	1058	896	147530	114594	223,9	25570
8	1054	901	148513	115577	225,8	25179
9	702	898	100270	115935	152,6	24689
10	694	891	100482	116147	153,2	24192
11	685	882	100583	116248	153,5	23692
12	674	873	100583	116247	153,7	23186
13	801	863	121451	119045	185,7	23637
14	790	853	121725	117175	186,4	22271
15	595	843	93630	116559	143,5	21693
16	571	826	92217	115146	141,4	21009
17	532	802	88853	111782	136,1	20055
18	445	763	78943	101872	120,5	18566
19	268	654	55254	74943	82,9	14102
20	46	477	3839	70606	5,60	13406
21	42	450	2345	69112	3,41	13303
22	26	87	8919	3187	12,36	2207
23	113	77	8249	2518	11,25	2078
24	104	74	7792	2060	10,52	1978
25	98	73	6992	1260	9,33	1941

Table A.5a Mass and vapor flow on each stage of preflash column (Base case)

Stage	Liquid flow on stage (kmol/hr):	Vapor flow on stage (kmol/hr):	Mass flow liquid from stage (kg/hr):	Mass flow vapor to stage (kg/hr):	Volume flow liquid from stage (m^3/hr):	Volume flow vapor to stage (m^3/hr):
1	1127	7,72	109721	110198	24752	2885436
2	1163	1175	126790	140081	29586	3146594
3	1221	1337	138223	151514	32641	3102258
4	1218	1394	141476	154767	33669	2934344
5	1195	1391	142350	155641	34102	2748435
6	1153	1368	141781	155072	34166	2559195
7	1075	1326	137951	151243	33398	2342691
8	877	1248	121582	134873	29451	1995138
9	452	1050	78335	91626	18584	1426428
10	386	131	72499	790	16535	94634

Table A.5b Mass and vapor flow on each stage of preflash column (Optimum case)

Stage	Liquid flow on stage (kmol/hr):	Vapor flow on stage (kmol/hr):	Mass flow liquid from stage (kg/hr):	Mass flow vapor to stage (kg/hr):	Volume flow liquid from stage (m^3/hr):	Volume flow vapor to stage (m^3/hr):
1	491	15,8	47626	48608	71,5	8775
2	442	546	47405	58400	73,1	9134
3	459	594	51099	62094	79,6	8880
4	458	611	52046	63041	81,6	8378
5	449	610	52109	63104	82,2	7847
6	435	601	51665	62661	81,9	7323
7	410	587	50428	61423	80,3	6780
8	363	562	47059	58054	75	6115
9	256	515	37220	48215	59	5345
10	407	135	74795	790	112	627

CHAPTER 5
CONCLUSION AND RECOMMENDATION

5.1 Conclusion

The influence of feed temperature and strippers' position on CDU profile was studied by employing variation on the furnace's temperature in preflash column and strippers' positions on pipestil column. The optimal operating conditions were obtained by calculating the maximum profit of USD54.93/m^3, it happened when the temperature in the furnace of preflash column was at 345°C while strippers′ position on pipestil was at S1p.5-4, S2p.14-13 and S3p.19-18. Finally the CDU profile was verified, the results of this research are as follow:

1. When the feed temperature increases, the column's temperature profile changes substantially and the appearance of a steep in tray above the feed tray. Since the temperatre on top of the column is constant, increasing the larger of the feed temperature that comes to the flash zone of preflash column cause a larger deviation of the temperature between the top and bottom of the column and this increases the amount of condensed liquid that is why the temperature in the bottom was high and it caused the steep to appear.

2. Incresing the feed temperature causes an increment of L/V, which means that fractionation is improved, however, the volume of liquid which descends along the column is greater and the internal reflux is improved which minimized the over vaporized materials without compromising the quality of product.

3. The pressure profile of both case in both columns are linear and identical to each other, it means that there are no influences from feed temperature or striper position on pressure profile of each columns tray.

4. Displacing kerosene stripper from p.6-5 to p.5-4 increases the rate of downflow liquid in overal column. This increment of downflow liquid causes an

increament of heat duty in the pipestil furnace because of the need of higher temperature to ensure vaporization of back liquid, that is why the temperature profile on the bottom trays was high.

5. Increasing the mass flow of the feed by 20%, the mass flow of the lights ends and heavy naphtha reached a new steady-state at higher and low rate flows; however, the mass flow of light naphtha and diesel flow rate unexpectedly, although slightly, decreased in both type of disturbance but the time necessary to reach a new steady-state for all product flow rate was at 1.5 hr. From this analysis of stream product flow the deviations was small that is why do not influence the products composition significantly.

6. No significant changes was found on CDU profile when the feed temperature or stripper position was changed, the deviation is normal. The column operation was satisfying for all cases no disturbance was verified in the CDU operation.

5.2 Recommendation

In this study the suggestions to be considered is comparison of data obtained with real plant in order to know the accuracy of those determined conditions. Need also to change the percentages of individual crude oil used to understand how it influence in the production of diesel and kerosene the more expensive products and higher demand.

RFERENCES

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1. Aspen Plus (2006), *Physical Property Methods and Models,* Aspen Plus Inc., Cambridge.
2. Bagajewics, M. (1997), "On the design flexibility of atmospheric crude fractionation units", *Chem. Eng. Comm.,* Vol. 166, pag. 111-136.
3. Bagajewics, M. and Ji, S. (2001), "Rigorous procedure for the design of conventional atmospheric crude fractionation units. Part 1: Targeting", *Ind. Eng. Chem.,* Vol. 40, pag. 617-626.
4. Budhiarto, A. (2009), *Buku Pintar Migas Indonesia*, Ditigem Migas-ESDM, Jakarta.
5. Direktorate Jenderal (2015), *Harga Minyak Mentah Indonesia untuk Bulan Augustus 2015,* Jakarta.
6. Errico, M. (2009), "Energy saving in a crude distillation unit by a preflash implementation", *Applied thermal engineering,* Vol. 29, pag. 1642-1647.
7. Fahim, M., Al-Sahhaf, T. and Elkilani, A. (2010), *Fundamentals of petroleum refining,* Elsevier, UK, pag. 83.
8. Gandalla, M. (2013), "A new optimization based retrofit approach for revamping an Egyptian crude oil distillation unit", *Proceedings of 13th conference – Advancements in renewable energy and clean environment,* Ed: Kamel, D. et al., British University in Egypt, Cairo, pag. 454-464.
9. Gutierrez, J. (2014), "Thermodynamic Properties for the Simulation of Crude Oil", *Jounal of engineering research and applications,* vol.4, pag.190-194.
10. Handogo, R. (2011), "Optimization on the performance of crude distillation unit (CDU)", *Asia-pacific Journal of Chemical Engineering*, Vol. 1, No. 7, pag. S78-S87.
11. Haydary, J. and Pavlik, T. (2009), "Steady-state aned dynamic simulation of crude oil distillation using Aspen plus and Aspen dynamic", *Petrolum and coal,* vol. 51, pag. 100-109.
12. Hossein, H.S., Zarei, A. and Rahimi, R. (2013), "CFD and experimental studies of liquid weeping in circular sieve tray columns", *Chemical Engineering Research and Design,* No. 91, pag. 2333-2345.

13. Jobson, M. (2014), *Distillations: Fundamentals and principles,* Elsevier Inc. Manchester.
14. Jones, D. S. and Pujadó, P.R. (2006), *Handbook of petroleum processing*, Springer, USA.
15. Kirgina, M.V. (2014), "Complex system form gasoline blending maintenance", *Proceeding of 15th international conference - Chemistry and Chemical engineering in XXI,* Eds. Sakhnevitch, B.V. et al., National Research Tomsk Polytechnic University, Russia, pag. 289-296.
16. Luo, Y. (2013), "Simultaneous optimization for heat-integrated crude oil distillation system", *Proceding of the 6th international conference on process systems engineering,* Eds: Wang, H. et al., Tianjin University, Tianjin, pag. 25-27.
17. Luyben, L. (2006), *Distillation Design and Control using Aspen Simulation*, New Jersey.
18. Malvin, A., Chan, A. and Lau, P.L. (2014), "CFD study of distillation sieve tray flow regimes using the droplets size distribution technique", *Journal of the Taiwan Institute of Chemical Engineers,* No. 45, pag. 1354-1368.
19. Muraza, O. and Ahmad, G. (2015), "Role of zeolite catalysts for benzene removal from gasoline via alkylation: A review", *Microporous and Mesoporous Materiales*, No. 213, pag. 169-180.
20. Oil Price Information Sistem (2016), a Dayly Report on Asia naphtha, LPG and Gasoline Spot Price, www.globalpetrolprices.com acess data February 16th 2016.
21. Puspitasari, E. and Setyarini, A. (2010), *Optimasi kinerja crude distillation unit*, Undergraduate final project, Institut Teknologi Sepuluh Nopember, Surabaya.
22. Rahimpour, M.R., Irashahi, D., Paymooni, K., Bahmanpour, A.M., and Shariati, A. (2011), "Simultaneous hydrogen and aromatics enhancement by obtaining optimum temperaturevbprofile and hydrogen removal in naphtha reforming; a novel theorical study", *International Journal Hydrogen Energy*, No. 36, pag. 8316-8326.

23. Ravindran, A., Ragsdell, K.M. and Reklaitis, G.V. (2006), *Engineering optimization, methods and applications*, John Wiley & Sons, Inc., New Jersey, pag. 2.

24. Saraf, D. (2010), “A crude distillation unit model suitable for online applications”, *Fuel processing technology*, vol. 73, pag. 1-21.

25. Seo, J.W., Oh, M. and Lee, T.H. (2000), “Design optimization of crude oil distillation”, *Chemical engineering technology,* Vol. 23, No. 3.

26. Silvestre, F. (2005), *Inferência da curva de destillação ASTM da destilação atmosférica para controle de avan,cado,* Undgraduate final project, Universidade federal de Santa Catarina, Florianópolis.

27. Simanzhenkov, V. and Idem, R. (2003), *Crude oil chemistry*, Marcel Denver Inc., New York.

28. Speight, G.S. (2011), *The refinery of future*, first edition, Elsevier, UK.

29. Speight, J. (2014), *High acid crudes*, Elsevier, USA

30. Viswanatan, J. and Grossman, I.E. (1993), Optimal feed locations and number of trays for distillation columns with multiple feeds”, *Engineering Industrial Design Center,* Carnegie Mellon University, No. 32, Pennsylvania, pag. 2942-2949.

31. Waheed, M.A. and Oni, A.O. (2015), “Perfomance improvement of a crude oil distillation unit”, *Aplied thermal engineering,* Vol. 75, pag. 315-324.

32. Wankat, P.C. (2015), “Decreasing costs of distillation columns with vapor feed”s, *Chemical Engineering Science,* No. 137, pag. 955-963.

33. Wauquier, J.P. (1998), *Separation process*, second ed,, TECHNIP, Paris.

34. Watkins, R.N. (1979), *Petroleum refinery distillation*, second edition, Gulf Publishing Company, Houston.

35. Yahaya, G.O., Hamad, F., Bahamdan, A., Tammanan, V.R and Hamad, E.Z. (2013), “Supported ionic liquid membrane and liquid-liquid extraction using membrane for removal of sulfur compounds from diesel/crude oil”, *Fuel processing technology*, No. 113, pag. 123-129.

36. YVC Rao (2004), *Chemical engineering thermodynamic,* India, pag. 232.

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Greetings,

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