MODELING, SIMULATION AND CONTROL OF A MODULAR REFINERY

by

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CERTIFICATION

This is to certify that the thesis

MODELING, SIMULATION AND CONTROL OF A MODULAR REFINERY

Submitted to the

Department of Chemical Engineering, University of Lagos

For the award of the degree of Bachelor of Science in Chemical Engineering is a record of original research carried out by

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The Head
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Dear Sir,

LETTER OF TRANSMITTAL

In accordance with the regulations of the Department of Chemical and Petroleum Engineering, University of Lagos, I hereby submit the research project titled MODELING, SIMULATION, AND CONTROL OF A MODULAR REFINERY in partial fulfilment of the requirements for the award of Bachelor of Science in Chemical Engineering at the University of Lagos.

Yours faithfully

Tobiloba Emmanuel OGUNLEYE

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DEDICATION

I dedicate this work to God Almighty who has been my strength and hope and my family for their unending prayers which propelled me towards the completion of this project.

ACKNOWLEDGEMENTS

I want to thank all who have contributed to the completion of this project in one way or another, I am indeed grateful.

Special thanks to my project supervisor, Dr. T.O Ajayi for her guidance throughout the course of this research project and most importantly for always correcting my errors with love. I also thank Walter & Smith modular refinery for providing the process data used in modeling this project.

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ABSTRACT

Modular refineries are refineries whose hardware are built in fragments with the intent of easy transportation, installation, and decommission. A modular refinery can produce a range of specific petroleum products, commonly naphtha, diesel, kerosene, and residue, at a time thereby meeting the immediate fuel needs of Nigeria. The establishment of modular refineries at strategic locations would reduce the importation of refined petroleum products, curb pipeline vandalism, impact the nation's GDP, and provide employment opportunities for both skilled and unskilled workforce. The daily consumption of diesel in Nigeria according to NMDPRA in 2021 was about 13 million litres per day, hence it is necessary to control the liquid level in the diesel chimney tray of the crude distillation column.

This work develops an input-output transfer function model used for designing the diesel chimney tray liquid level in the crude distillation column. Four tuning methods were used: Ziegler Nicholas; Cohen-Coon; Integral Time Weighted Absolute Error (ITAE); and Internal Model Control (IMC) tuning methods. The ITAE tuning method with controller parameters KC, τi , and τD of 2.48154, 2.14899, and 0.31699 respectively gave the best dynamic response with 4% overshoot, 2.5s rise time, 7s settling time, and minimal oscillation.

Tuning the programmable logic controller in the distributed control system of Walter-Smith modular refinery using the ITAE process parameters gotten above would minimize weeping and flooding in the column thereby ensuring maximum draw of product and maintaining temperature and pressure distribution across the tray, thus ensuring premium product quality.

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NOTATIONS

MESH	Material, Equilibrium, Summation and Heat
CDU	Crude Distillation Unit
FC	Flow Control
PC	Pressure Control
TC	Temperature Control
LC	Level Control
H_N	Liquid Hold-up on Tray N
Y_N	Vapor Composition on Tray N
L_{N+1}	Liquid Flowrate on Tray N-1
C_p	Specific Heat Capacity
S_G	Specific Gravity
T	Temperature
F	Feed Flowrate
ρ	Density
P	Proportional Only Controller
PI	Proportional-Integral Controller
PID	Proportional-Integral-Derivative Controller
ZN	Ziegler-Nicholas
K_c	Proportional Gain
τ_i	Integral Time
τ_d	Derivative Time
ITAE	Integral Time Weighted Absolute Error
IMC	Internal Model Control
K	Effective Gain
t_d	Time Delay
τ	Time Constant

CHAPTER ONE - INTRODUCTION

1.1 Background of Study

Over the years, the human race has had several opinions about its origin narrowed down to three ideas namely: evolution, creation by God, and unidentified flying objects (UFO's). Scientists believe the first humans lived during the old stone age (about 3.5 million years ago) where they moved from place to place in search of food and shelter and competed with other animals for the natural products of the environment in which they lived. They were physically defenseless, and hence served as food to wild animals and were also prone to destruction by natural disasters as they slept in trees or huddled themselves on the ground (Wrigley, 2015).

In an attempt to survive and adapt to this situation, the early humans made tools from stone for defense and hunting, mastered the use of fire for cooking and warmth in cold seasons, lived in caves for safe shelter and most importantly developed oral language. The population of man was declining as war and disease resulted in high mortality rates, especially among children (Gowlett, 2016).

Over time, as man increased in intelligence, he ventured into subsistence farming, domestication of animals for food & transportation, invention of metal weapons & traps for hunting, mastering of pottery for food storage, cloths and jewelry, construction of boats and nets for fishing etc. This age was called the New Stone Age. Mortality rate was greatly reduced and life expectancy increased due to improved living conditions leading to increase in population (Potts, 2000). As man's population increased, the rate of production could no longer meet the demand for food and other basic needs which in turn made the prices of commodities skyrocket. This led to the

“Agricultural Revolution” of the late 17th and early 18th century that set the stage for industrialization (Elliot, 2016).

The discovery of coal, oil and gas (non-renewable fossil fuels) played a major role in industrialization. The first industrial revolution started in Britain (1820’s – 1840’s) and was characterized by major advancements in the textile industry. It was driven by the invention of the steam engine that runs on coal (the most abundant fossil fuel), which made it possible for machines to effectively do the work previously done by humans and animals. Despite its abundance on earth, coal is not a perfect fuel as it released substantial amount of impurities (Sulphur and nitrogen) into the atmosphere on combustion which caused poisoning and acid rain (Syed, 2014).

The search for a cleaner source of energy led to the second industrial revolution characterized by the discovery of oil between the late 19th and early 20th century. One of the driving forces behind the increased demand in crude oil was the invention of a kerosene lamp. Kerosene, a petroleum product was in high demand because it was a cleaner-burning and more-reliable fuel for lighting. Decades later, automobiles with gasoline-burning engines became conventional and the need for petroleum products became evident once again. The growth of the automobile industry continued to stimulate the oil industry, particularly after World War II which made oil consumption grow at a rate of 7% per annum (Morgan, 2017).

Petroleum also known as crude oil (the most widely used fossil fuel in the world) is a naturally occurring liquid mixture of thousands of compounds. Hundreds of different crude oils (usually identified by origin) are continually processed in the world's refineries. Most of the compounds in crude oil are hydrocarbons (organic compounds composed of carbon and hydrogen atoms). Other compounds in crude oil contain not only carbon and hydrogen, but also small (but important) amounts of other elements –most notably sulfur, as well as nitrogen and certain metals (e.g., nickel, vanadium, etc.). The compounds that make up crude oil range from the smallest and simplest hydrocarbon molecule – CH_4 (methane) – to large, complex molecules containing up to 50 or more carbon atoms (as well hydrogen and hetero-elements) (National Programme on Technology Enhanced Learning, 2006).

In refineries, crude oil can be converted into more than 2500 refined products, including liquefied petroleum gas, gasoline, kerosene, aviation fuel, diesel fuel, fuel oils, lubricating oils, and feedstock for the petrochemical industry. The transformation, both physical and chemical, that crude oil undergoes takes place in separate processing units.

Conventional petroleum refineries are large industrial structures comprising of different crude processing units arranged in series. These processing units and their associated operations are:

i. Separation processes: Atmospheric distillation, Vacuum distillation, and Light ends recovery (gas processing);

ii. Petroleum conversion processes: Cracking (thermal and catalytic), Reforming, Alkylation, Polymerization, Isomerization, Coking, and Vis-breaking;

iii. Petroleum treating processes: hydro-desulfurization, Hydro-treating, Chemical sweetening, Acid gas removal, and De-asphalting (National Programme on Technology Enhanced Learning, 2006)

When any one of these processing units are built in modules, they are referred to as modular refineries. Conventional refineries can process up to 100,000 barrels per day (bpd) and above while modular refineries can only process 500 to 30,000 bpd (Angela *et al.*, 2019)

Jim Smith Hogan was the first to develop and patent a one skid crude topping unit oil refinery (Fig 1.1). The invention was driven by the need to save cost of moving crude oil from a remote reservoir to refineries and the fact that it can be hidden or dismantled if the need arises. The skid was 12 feet wide, 45 feet long and 11 feet high and could be transported by a tow truck trailer, sea transport container or a railway container. This mini refinery could process about 750 to 1500 barrels of crude oil per day to produce gasoline, diesel oil and heavy fuel oil residue. The only problem with this invention was that the equipment were crowded on the module which made interconnection with other skid mounted units difficult (Hogan, 1976)

Over the years, the production of modular topping units on one skid was phased out and multiple skid units were employed. They can be mounted on a concrete foundation horizontally, vertically, single level or multi-level depending on available land space and interconnected in the field with other skid mounted components to create a small remote refining unit or linked together in a cascade to form a large refinery (Mamudu, Igwe and Okonkwo, 2019).

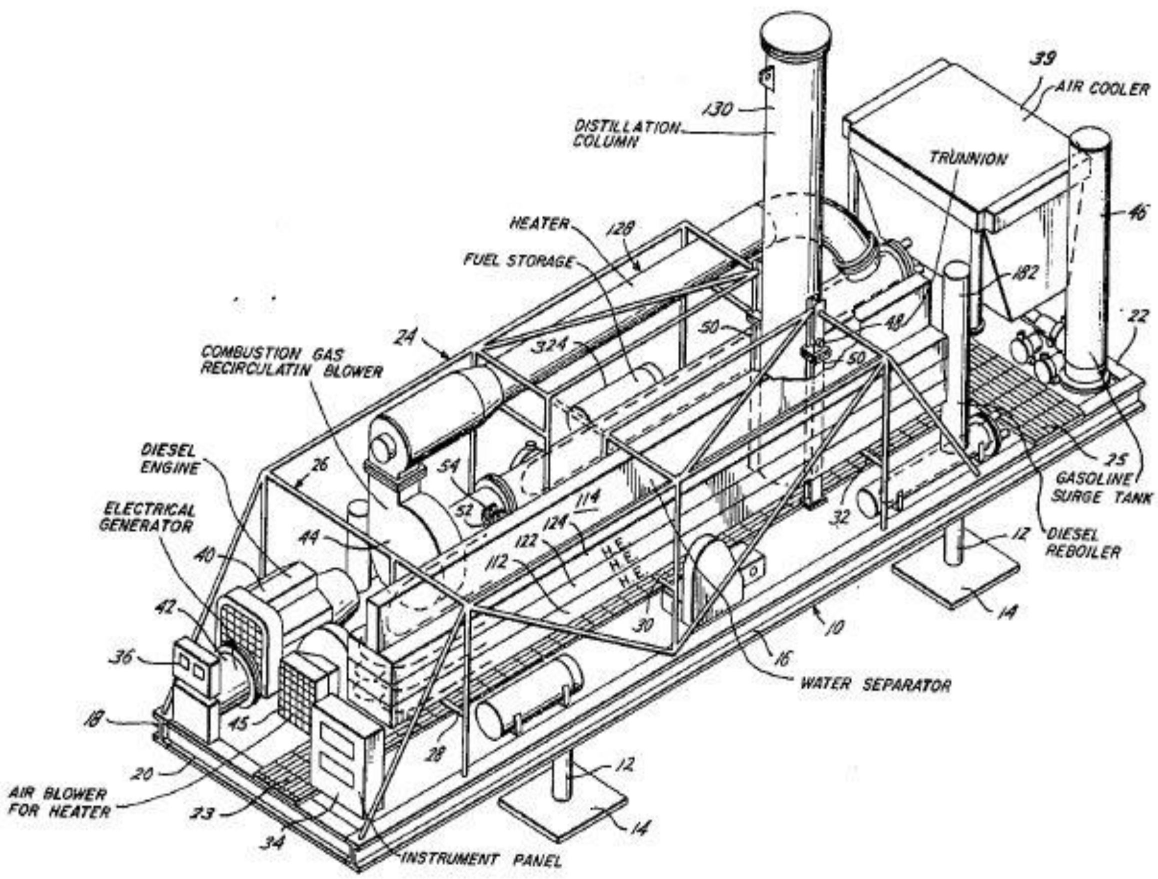

Figure 1. 1 The First Modular Refinery (Hogan, 1976)

There are 42 refineries in Africa with a total capacity of 3,217,600 barrels per day (BPD). The major oil producing countries are Egypt with nine refineries (774,900 bpd), Algeria with five refineries (303,700bpd), Libya with five refineries (380,000 bpd), South Africa with four refineries (545,000 bpd) and Nigeria with 4 refineries (445,000 bpd).

Nigeria, a member of the Organization of Petroleum Exporting Countries (OPEC) has been one of the world's leading importer of refined petroleum products for over twenty years now. Although Nigeria presently has four conventional refineries (Kaduna, Warri, Old Port-Harcourt and the New Port- Harcourt) with a combined operating capacity of 445,000 bpd, it currently produces less than

100,000 bpd which accounts for about 18-22% of its total capacity (Ogbon, Otanocha and Rim-Rukeh, 2018)

This low output capacity has been attributed to inadequate funding and autonomy, bad maintenance culture, political instability, poor management, irregular feedstock supplies, non-implementation of the Petroleum Industry Bill (PIB), ineffective technical services department, pipeline vandalism, use of obsolete technologies and delayed turn around maintenance.

In a bid to overcome the problem of inefficiency in Nigerian refineries, the concept of modularization should be embraced. Modularization can be defined as splitting of a system into smaller units with distinct functions to improve the overall efficiency of the whole unit.

1.2 Problem Statement

Despite the fact that Nigeria has the tenth largest crude oil reserve, they have been one of the leading importers of refined petroleum products for over twenty years. Out of the combined operating capacity of 445,000 BPD (Barrels per day) from its four refineries, Nigeria produces less than 100,000 BPD and has recently hit an abysmal output capacity in recent times. The low output capacity has been attributed to inefficiency of the conventional refineries.

In an attempt to provide a solution to the problems stated above, the nation was left with the following options: establishment of more conventional refineries, legalization of illicit refineries, swapping of raw petroleum with refined products from different nations and establishment of modular refineries around the nation. After much debate, the most plausible alternative was the construction of modular refineries at specific areas within the nation.

1.3 Justification of Project

The low output capacity and inefficiency of the conventional refineries created an avenue for gross mismanagement of funds and corruption as it costs the country 10.23 billion naira in operational and administrative expenses despite processing no crude in 2020.

This indicates the need for the establishment, optimization, and control of modular refineries which offer the advantages of flexibility, simpler maintenance procedures, and so on over conventional refineries.

1.4 Aims and Objectives of Study

The aim of this project is to understand the workings of a modular refinery by developing a transfer function model from first principles to be used for control purposes while the objective is to use the model developed to design and tune a PID controller with various tuning methods so as to determine the tuning method which gives the best dynamic response.

1.5 Relevance of the Study

The establishment of modular refineries at strategic locations like remote areas of the Niger-Delta, would reduce the importation of refined petroleum products, pipeline vandalism, environmental pollution, mismanagement of funds and corruption in the oil & gas industry.

It would also reduce the importation of petroleum products, attract foreign direct investments (FDI) into the oil & gas industry thereby creating more job opportunities in the oil sector and improve the Nigerian standard of living.

CHAPTER TWO - LITERATURE REVIEW

For nearly a century, chemical engineering has been governed by two fundamental tenets: economy of scale and unit operations. The first dictates that increasing the capacity of chemical plants will make their construction more capital-efficient and improve the utilization of available resources while the second is based on the fact that all plants are designed and built using a relatively uniform set of building blocks i.e unit operations. In the past two decades, these rules were challenged by some who aimed to alter the conventional "one unit- one operation" approach by splitting the same plant into multiple unit operations, a concept known as process intensification (Baldea *et al.*, 2017).

These intensified systems were smaller in size, and more efficient in terms of transportation, safety, energy consumption, and environmental performance when compared to their conventional counterpart. The only limitation was low production capacity which could be overcome by increasing the number of intensified systems operating in parallel.

2.1 Modularization

A module can be defined as a unit or a set of individual units which when integrated with other units can form a complete system. This definition immediately brings to mind the concept of process intensification however, modularity goes beyond the scope of unit operations as it allows a combination of other fundamental tasks (e.g chemical reaction in unit processes) in the same unit. Modularization can therefore be defined as the splitting of a system into smaller units with distinct functions to improve overall efficiency. While modularization is not new in chemical process industries, it is not so prevalent in other manufacturing sector (Baldea *et al.*, 2017).\

2.2 Modular Refinery

Cenam Energy Partners (2014) described a modular refinery as a refinery whose hardware is built in modules intended to be easily moved to any place in the world in a variety of sizes with limit's that go from 500 to 30,000 barrels per day. (Mamudu *et al.*, 2016) (as cited in SHUMWAY, RICH and BROWN, 2003) also characterized a modular refinery as a regular refinery built in a fragmented manner, or just a major refinery in a scaled down structure (Angela *et al.*, 2016).

The crude distillation unit is the basic unit of a modular refinery as it produces intermediate products that can serve as feedstock for subsequent downstream units Yavini, Namo and Sani, 2015). Although there are various configurations of a crude distillation unit, it mainly consists of a distillation column, condenser, heat exchanger networks and a furnace while other features remain optional (Angela *et al.*, 2016).

Crude oil contains salts, water and contaminant sediments which can be harmful to downstream equipment and must therefore be removed. To remove the salts, water (about 3 – 10% of the crude oil volume) is mixed with the crude oil and typically heated to temperatures between about 215˚F to about 280˚F and allowed to separate in the de-salter (Idris *et al.*, 2018). The desalted crude enters another heat exchanger network called the pre-heating train which make use of the vapors of the main column condenser, the pump-around circuit streams, and the products that need to be cooled. In the preheating train, the crude is under pressure to suppress vaporization (Mamudu *et al.*, 2016). The preheated crude then enters the furnace, where it is heated to about 340-370 °C. The partially vaporized crude is fed into the flash zone of the atmospheric column, where the vapor and liquid separate. The vapor includes all of the components that comprise the products, while the liquid is the residue with a small amount of relatively light components in the range of gas oil

(Bagajewicz and Ji, 2001). These components are removed from the residue by steam stripping at the bottom of the column and are either used as fuel oil or charged to a vacuum distillation unit. In addition to the overhead condenser, there are several pump-around circuits along the column, where liquid streams are withdrawn, cooled, and sent back to upper trays.

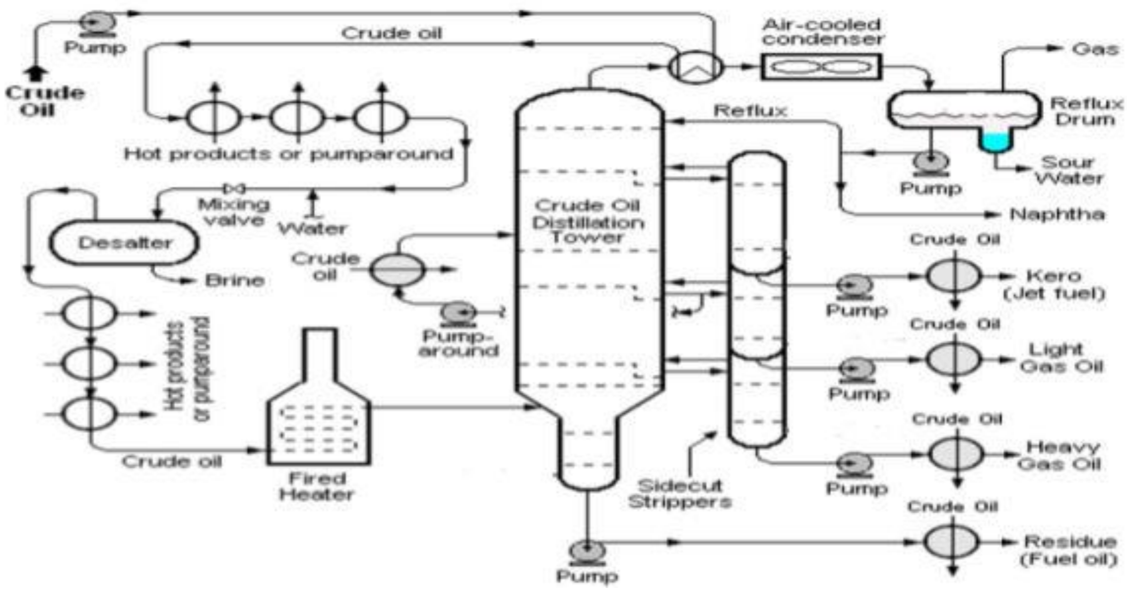

Figure 2. 1 A Modular Crude Distillation Unit (Mamudu et al., 2016)

Products such as heavy naphtha, kerosene and gas oil (diesel) are also withdrawn in the liquid state from different trays and then stripped by steam in side strippers to remove light components. Cooling water and sometimes air coolers are used in the heat exchangers, but it is always more advantageous to have these streams release their heat to the raw crude oil in the heat exchanger networks (pre-heating trains) (Idris *et al.*, 2018).

The subsequent sections of this chapter give a detailed literature review of the advantages & disadvantages, location, environmental limitations, optimum configuration, modeling & simulation, optimization studies and control of modular refineries.

2.2.1 Advantages and Disadvantages of a Modular Refinery

The idea of a modular refinery is not novel. Nigerians began to take interest in modular refineries when the government made it a major initiative in the quest to reposition the petroleum sector (Content, 2018). Although the establishment of modular refineries has been driven by the need for a more flexible production (in terms of capacity, product type and geographical location), it has its own advantages and disadvantages when compared to a conventional refinery (Ogbon, Otanocha and Rim-Rukeh, 2018).

The advantages of a modular refinery when compared to a conventional refinery are ease and faster speed of construction, lower costs (capital expenditure, installation, operation), quicker investment recovery, production of specific products at a time, use of limited land space, flexibility to meet changes in demand, ease of scalability, lower energy consumption, reduced CO_2 emissions, lower workforce, simplified maintenance, lower downtime and better safety and control practices (Baldea *et al.*, 2017) .

The only disadvantage of a modular refinery when compared to a conventional refinery is its lower production capacity and this can be overcome by setting up various modular refineries at strategic locations within the country. The table 2.1 below lists the juxtaposes a modular and conventional refinery based on certain factors.

Table 2. 1 Modular Refineries vs Conventional Refineries

Factor of Comparison	Modular	Conventional
Production capacity	4,000 to 50,000 bpd	120,000 bpd and above
Time to set up	12 – 18 months	2 – 5 years
Flexibility	Can adapt to meet changes in demand	Complicated
Range of Products	Caters for just some petroleum products which include unleaded gasoline, kerosene, diesel, and fuel oil	Caters for all range of petroleum products
Construction process	Constructed in modules for easy transportation	Constructed on site
Land size	Small land space required	Large land space required
Installation cost	Low	High
Return on investment	Rapid	Slow

2.2.2 Location of a Modular Refinery

The process of selecting the location for the establishment of a modular refinery is just as important as its design and optimization. Modular refineries should therefore be set up at strategic locations as they directly affect the economics, profitability & efficiency of crude oil refining especially at start-up (DPR , 2017). The factors that affect the location of a modular refinery include; closeness to source of crude oil (i.e oil producing marginal fields & flows stations), product market outlet, proximity to industrial hubs, political and environmental factors etc. (Adewuyi, 2017). It is a general rule of thumb that logistics costs can be greatly reduced by moving production closer to source of raw materials and/or customers (Baldea *et al.*, 2017).

(Adewuyi, 2017) conducted a research to determine how the location of a flow station from which six modular refineries (the refineries have equal capacities of 20,000bpd sited at different locations with regular intervals of 10km from the flow station) receive feedstock affect the pressure drop along the pipe lines and hence the capital cost of keeping the chemical composition of crude oil being transported and operating conditions of the network constant. He then developed an equation to show the relationship between pressure drop (ΔP, bar/100m) along a pipeline as a function of pipeline length (l, km) (Equation 2.31) and also that between capital cost (USD) needed to establish a modular refinery as a function of pipeline length (Equation 2.32). From the equations (2.31 – 2.32), it was concluded that the farther the distance between a modular refinery and it's source of crude oil, the higher the pressure drop and hence the higher the capital costs since pumps would be installed along the pipelines to compensate for pressure drop.

$\Delta P = 1.6341\, l + 0.3212$……………. (Equation 2.31)

$\Delta C = 1{,}773{,}881\, l + 880{,}000$ ……… (Equation 2.32)

The location of a modular refinery also depends on policies and environmental regulations put in place by the government and international bodies to curb pollution which is mainly in form of CO_2 emissions from burning fossil fuels used in the refinery.

2.2.3 Environmental Limitations of a Modular Refinery

Carbon dioxide (CO_2) as a greenhouse gas plays a vital role in global warming. Studies show that it is responsible for about two thirds of the enhanced greenhouse effect (Gadalla *et al.*, 2005). To meet the environmental regulations as agreed in the Kyoto Protocol, the chemical process industries have been challenged to reduce their greenhouse emissions, particularly the release of CO_2 into the atmosphere (Gadalla *et al.*, 2006).

Globally, the petroleum refining sector ranks third among stationary CO_2 producers, after the power production sector and the cement industry. Other large producers are the iron and steel industry, and the petrochemical industry (Straelen *et al.*, 2009). Together, stationary sources amount to about 60% of overall global CO2 emissions. CO_2 emissions from refineries account for about 4% of the global CO_2 emissions which is close to 1 billion tons of CO_2 per year (Straelen *et al.*, 2009).

A refinery may use about 1.5% up to 8% of feed as fuel, depending on the complexity of the refinery. For a typical world-scale 300,000 barrel per day refinery, this will lead to CO_2 emissions ranging from 0.8 to 4.2 million tons of CO_2 per year (Al-Salem, 2015). CO_2 can be emitted from various sources in a conventional refinery which include; furnace & boilers (used to increase temperature of input materials for distillation and other unit operations), utilities (for production of electricity and steam), fluid catalytic cracker unit and hydrogen manufacturing unit (Straelen *et al.*, 2009).

In a modular refinery, CO_2 emissions are mainly from the furnaces, boilers & utilities and strongly depends on the type of fuel being burned (Al-mayyahi *et al.*, 2011). To reduce the CO_2 emissions from these devices, the first and most economical route is efficient energy conservation & heat recovery by process optimization. Other options include optimum configuration, the use of biofuels instead of fossil fuels, renewable sources of energy (wind or solar) for the production of heat and electricity, and carbon capture & storage (CCS) which can be oxy-firing, pre-combustion capture or post-combustion capture (Wanders, 2016).

(Gadalla *et al.*, 2005) proposed a model equation to calculate the amount of CO_2 emitted in a modular refinery. He then optimized the operating conditions (temperature of the feed, reflux ratio, stripping steam flow rates, temperature difference of each pump around, and the flow rate of the liquid through each pump around) of the crude distillation unit (CDU) to decrease CO_2 emissions and energy consumption. He also considered the integration of a gas turbine with the furnace to meet the optimization objectives. The integration of gas turbines was found to be more effective in reducing CO_2 emission.

(Al-mayyahi *et al.*, 2011) investigated the effect of two different blends of crude oil on energy consumption and consequently CO_2 emissions in a refinery. A rigorous model which incorporates pinch analysis was used to estimate the amount of CO_2 emitted in the CDU. It was observed that heavier crude blends recorded higher CO_2 emissions compared with the lighter blends and operating at maximum over-flash by maximizing the furnace outlet temperature was preferable for maximizing net revenue & reducing CO_2 emissions.

(Wanders, 2016) conducted an experiment to determine the total amount of CO_2 emitted from a modular refinery and found that the atmospheric distillation unit accounted for 70% of the total emissions. He applied a multi-criteria analysis to determine the most promising CO_2 reduction alternative. It was concluded that processing light & sweet crude oil was the most promising alternative for reducing CO_2 emissions in modular refineries followed by exchanging heat to nearby residential areas and the optimization of the atmospheric distillation unit.

(Gadalla *et al.*, 2015) also modeled an existing crude distillation unit and applied optimization and revamping techniques to minimize energy consumption and reduce CO_2 emissions. The three optimum configuration options considered were; the optimization of pump-around and heat exchanger network, installation of a pre-flash drum, and the addition of a third pump-around. They discovered that installation of pre-flash drums was the most effective in reducing CO_2 emissions.

2.2.4 Optimum Configuration of a Modular Refinery

The configuration of a modular refinery depends on the type of crude being refined, product specification, product yield, operating conditions of the refinery, environmental limitations and regulations, government policies, profitability, and so on. Since these factors differ from one location to another, there are different configurations for a modular refinery and not one can be referred to as the perfect configuration.

In the past decades, multiple researchers have proposed and developed various methods for the configuration of modular refineries. Conventional methods applied heuristic rules, experience, empirical correlations and simple relationships between variables to describe the system. A simple modular refinery (Figure 2.2) consists of an atmospheric distillation unit, a heat exchanger network, and a furnace as its main equipment. Products from this simplified version did not meet desired specifications since they were not stripped of any light components. To overcome this limitation, Miller and Osborne proposed the introduction of side strippers which used either steam or distilled hot residue oil as a source of heating (Miller and Osborne, 1938).

(Brugma, 1941) suggested the addition of pre-flash drums to avoid unnecessary heating of light naphtha which are condensed separately (Figure 2.3). (Nelson, 1958) estimated the number of trays in each section of the column and the stripping steam flow based on empirical correlations obtained from previously established designs.

(Watkins, 1979) also selected the number of trays in each section of a column from a predetermined range and estimated the stripping steam flow based on product flow rates. He reviewed Miller & Osborne's work and found that although their design was feasible, increasing vapor and liquid traffic will occur within the column since all heat within the system is handled only by the condenser. To overcome this limitation, Watkins made an improvement by removing some of the heat below the top tray, which led to the introduction of pump around and pump back refluxes (Figure 2.4).

The approach of the researchers mentioned above formed the basis for many subsequent design methodologies for modular refineries which involved iterations and trial & error procedures. Other researchers therefore focused on developing integrated design methods that address the design of the atmospheric distillation unit and the associated heat recovery network simultaneously which was omitted in previous designs.

(Golden, 1997) reviewed Brugma's work and proved that the light components removed helps to increase separation rate of atmospheric gas oil from the residue. He noticed that when the operating temperature of pre-flash drums was increased above a particular temperature, the atmospheric column distillate yield losses also increased. He therefore provided useful insights on how flash temperature and flash vapor feed location affect the performance of the crude distillation column.

(Liebmann, Dhole and Jobson, 1998) combined rigorous column models of Watkins (1979) and pinch analysis to design a modular refinery. His approach takes design decisions in a sequential manner considering heat recovery at each step using pinch analysis. He noticed that his indirect sequence of columns saved 20% of utility cost, reduced vapor flow rate and also observed that using reboilers rather than, or in conjunction with steam stripping improves separation efficiency but reduces heat recovery opportunity.

(Suphanit and Smith, 1999) applied the column decomposition strategy of Liebmann *et al* with Fenske-Underwood-Gilliand (FUG) shortcut model for the crude oil distillation column. He noticed that the FUG method underestimates the vapor flow in each column section and therefore proposed that enthalpy balances be carried out at the top and bottom of the column to adjust the minimum vapor flow rate. This increased the heat recovery opportunity and minimized the total annual cost.

(Sharma *et al.*, 1999) then proposed to use the concept of a column grand composite curve which identifies the maximum amount of energy that can be recovered from pump-arounds without affecting the side draw separation. They discovered that this approach eliminates the need for repeated column simulation thereby significantly saving design time. A limitation of this approach is that the role of the stripping steam is neglected.

(Bagajewicz and Ji, 2001) focused on overcoming the above limitation by incorporating the effect of the stripping steam on the maximum heat recovery of the crude distillation column and introducing the concept of a heat demand–supply diagram. The basis of this design was the transfer of heat from the condenser to the lower pump-around. They found out that when processing light crude, three pump-around are needed to utilize the heat surplus from the condenser, two pump-around are needed for an intermediate crude while an heavy crude could make do with only one pump-around as there is no substantial heat surplus in the region of the first pump-around.

(Ji and Bagajewicz, 2002) then presented a rigorous approach for setting the inlet and outlet stream conditions when designing a crude oil distillation unit with a pre-flash unit. They observed that at a pre-flash vapor temperature of 163°C and vapor feed tray location close to the top of the column increases the output of the distillate but not necessarily its yield. Thus, pre-flash drums can only be advantageous from the point of view of energy consumption if yield losses of atmospheric gas oil can be accepted.

(Rastogi, 2006) extended the shortcut model of Suphanit to account for column pressure drop and pump-around location. A detailed model of both the heat exchanger network and the distillation column was incorporated into an optimization framework that optimized the column structure and operating conditions. He concluded that the location of the pump-around can increase heat recovery opportunity and saves 23.9% of the total operating costs.

(Chen, 2008) modified the shortcut model of Rastogi to allow for other pump-around locations and also modeled temperature-dependent properties (heat capacity) of process streams undergoing phase change. He discovered that the location of pump-around combined with varying heat capacities of crude oil and other process streams saves energy by only 18% compared to the 31% of previous studies when heat capacities were assumed to be constant with temperature.

(Errico, Tola and Mascia, 2009) studied the effect of the pre-flash drum scheme which separates some light components or vapor before crude is passed through the furnace and compared it to that of the pre-fractionation column. They found out that the pre-fractionation column increases light distillate and reduces furnace load but does not reduce energy consumption as much as the pre-flash drum does. Therefore, one's choice would greatly depend on refining objectives.

(Wang *et al.*, 2011) applied thermodynamic metrics to select the best pre-separation scheme for heavy crude oils. Nine predistillation arrangements are explored, and the option with two pre-flash units was found to have performed best (Figure 2.3). Heat recovery is not explicitly considered, so the results do not relate directly to demand for fired heating; in addition, product quality specifications appear to be only partially addressed, via stream or ‘cut’ temperatures.

(Benali, Tondeur and Jaubert, 2012) demonstrated that pre-flash units can bring benefits in terms of exergy destruction, but not saving as much energy as claimed in previous literature. They claimed the fact that exergy destruction decreases by 14% and by more than 21% in the furnace alone, and energy consumption decreases by 19.3%, is not entirely true as the pre-flash drum was assumed to be operating at the same temperature as outlet stream (that is the two preheat trains are receiving the same amount of heat).

(Al-mayyahi, Hoadley and Rangaiah, 2014) utilized multi-objective optimization techniques to study the effects of single and multiple pre-flash units on both energy consumption and yield. Their study investigates the vapor feed location and considers heat integration with and without pre-flash units. They observed that two pre-flash drums in series (Figure 2.4) saved more energy than the single pre-flash drum and also reduces CO2 emissions for a high residue yield.

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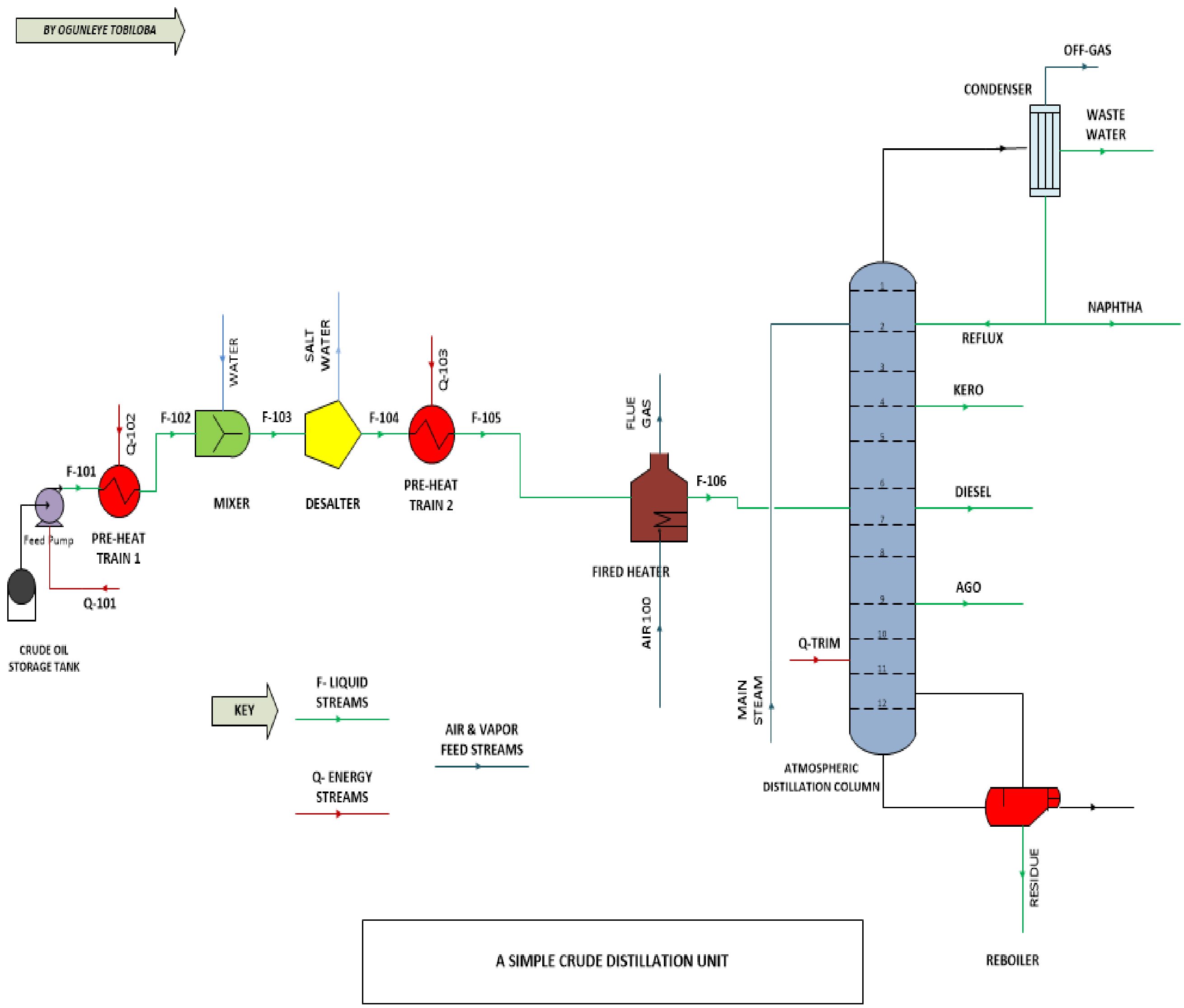


Figure 2. 2 Process Flow Diagram (PFD) of a Simple Crude Distillation Unit (CDU)

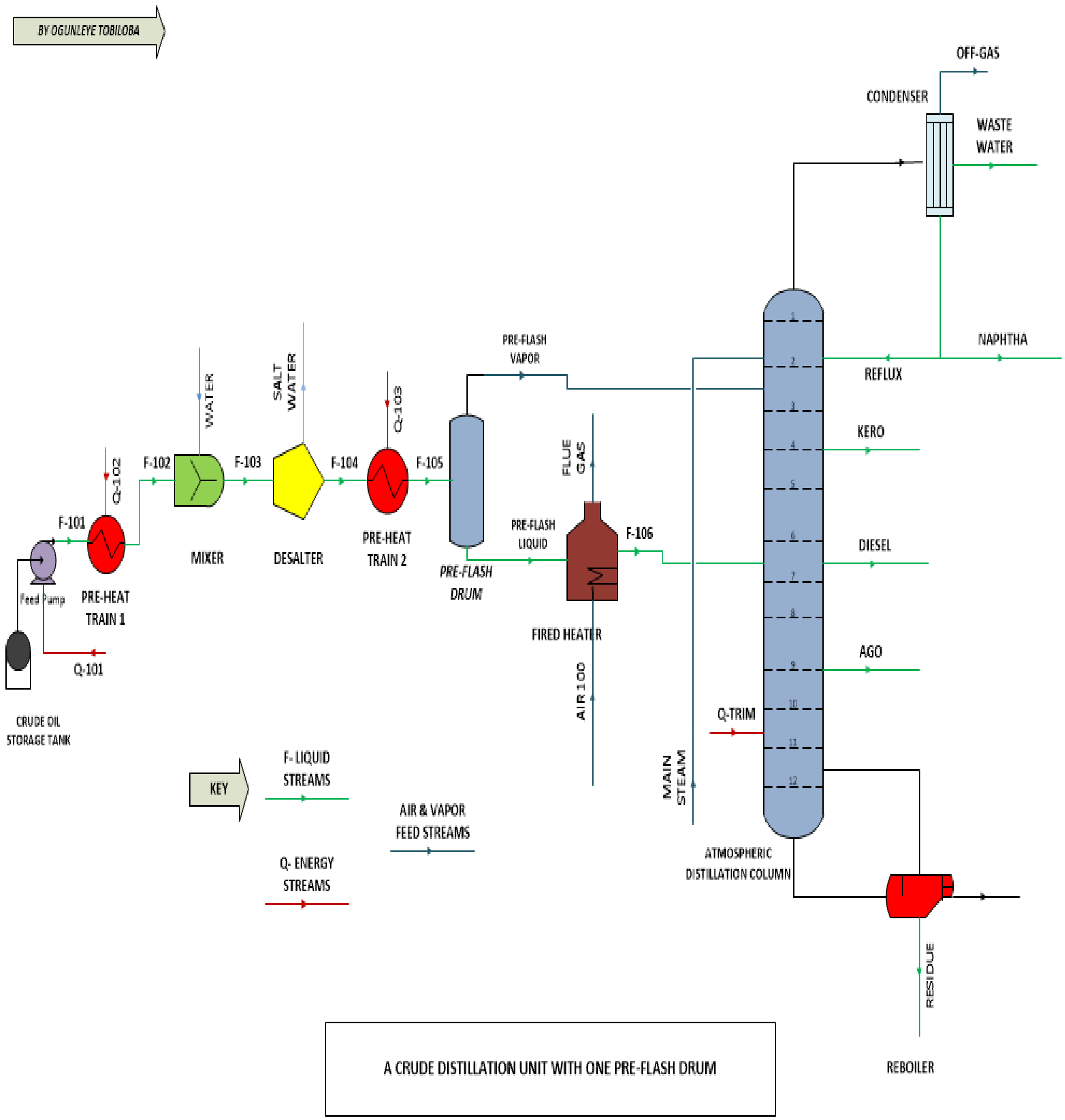


Figure 2. 3 PFD of a CDU with 1 pre-flash drum

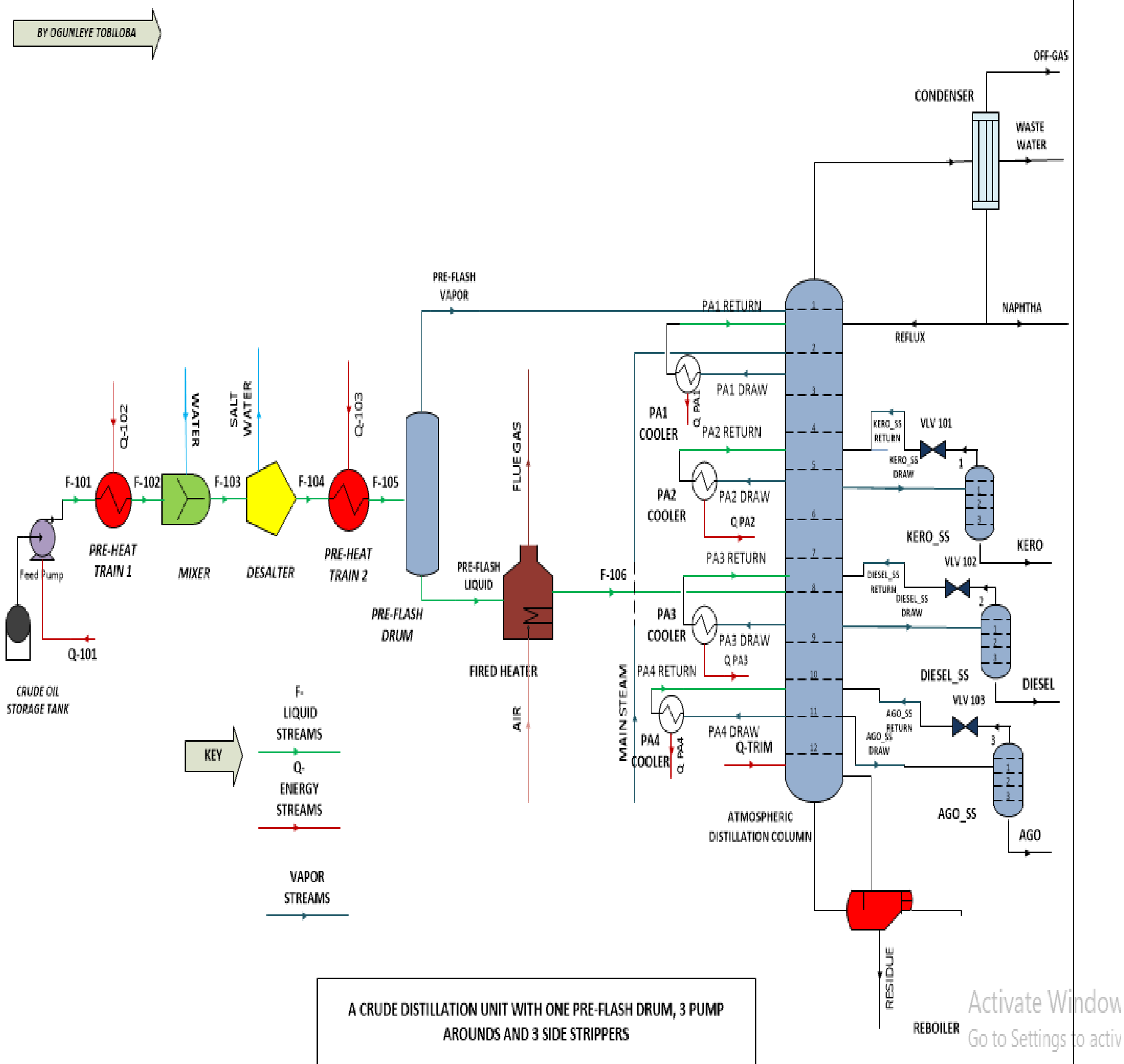


Figure 2. 4 PFD of a CDU with 1 Pre-flash drum, 3 Pump-around and 3 Side Stripper

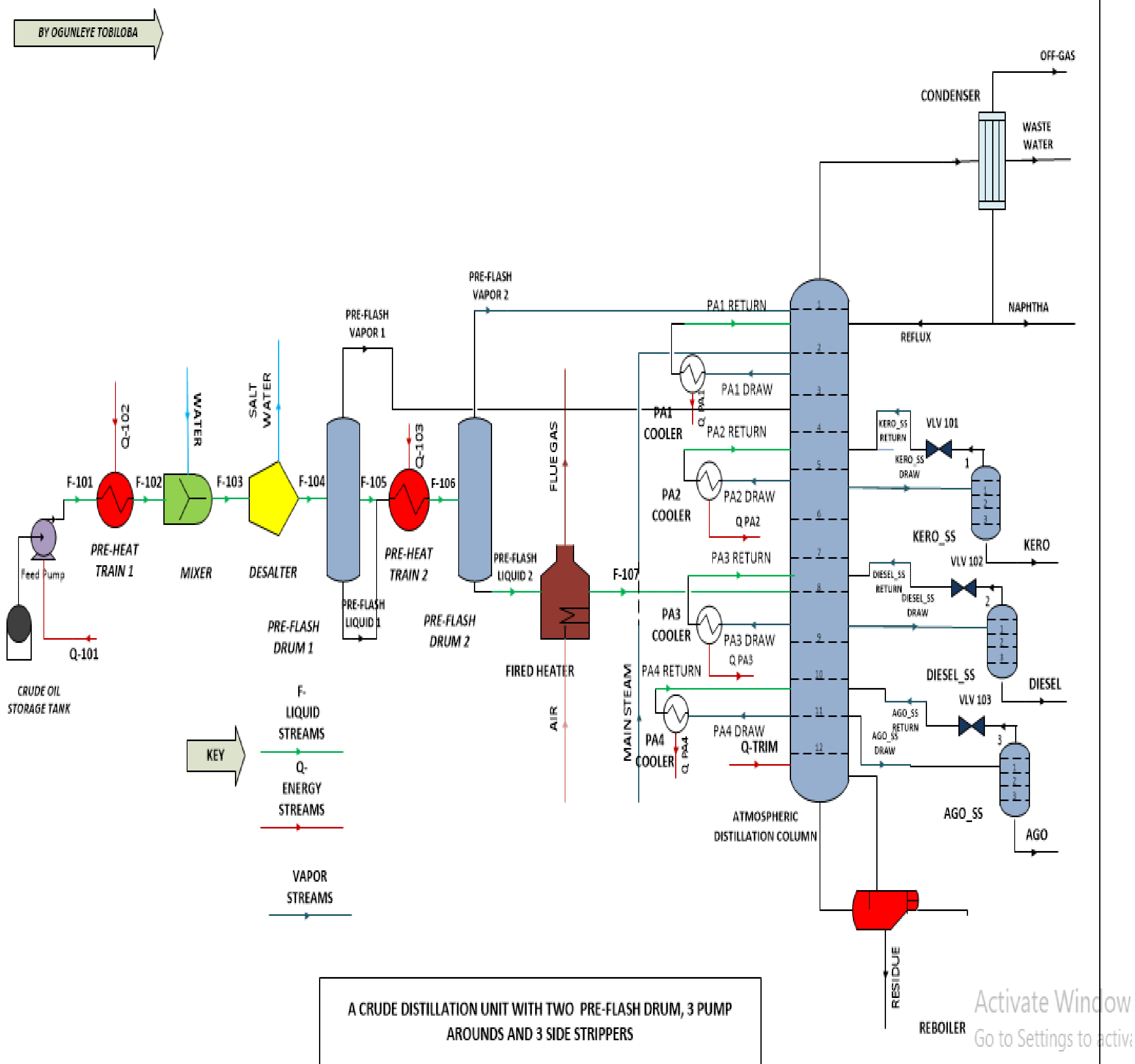


Figure 2. 5 PFD of a CDU with 2 Pre-flash Drums, 3 Pump-around, and 3 Side Stripper

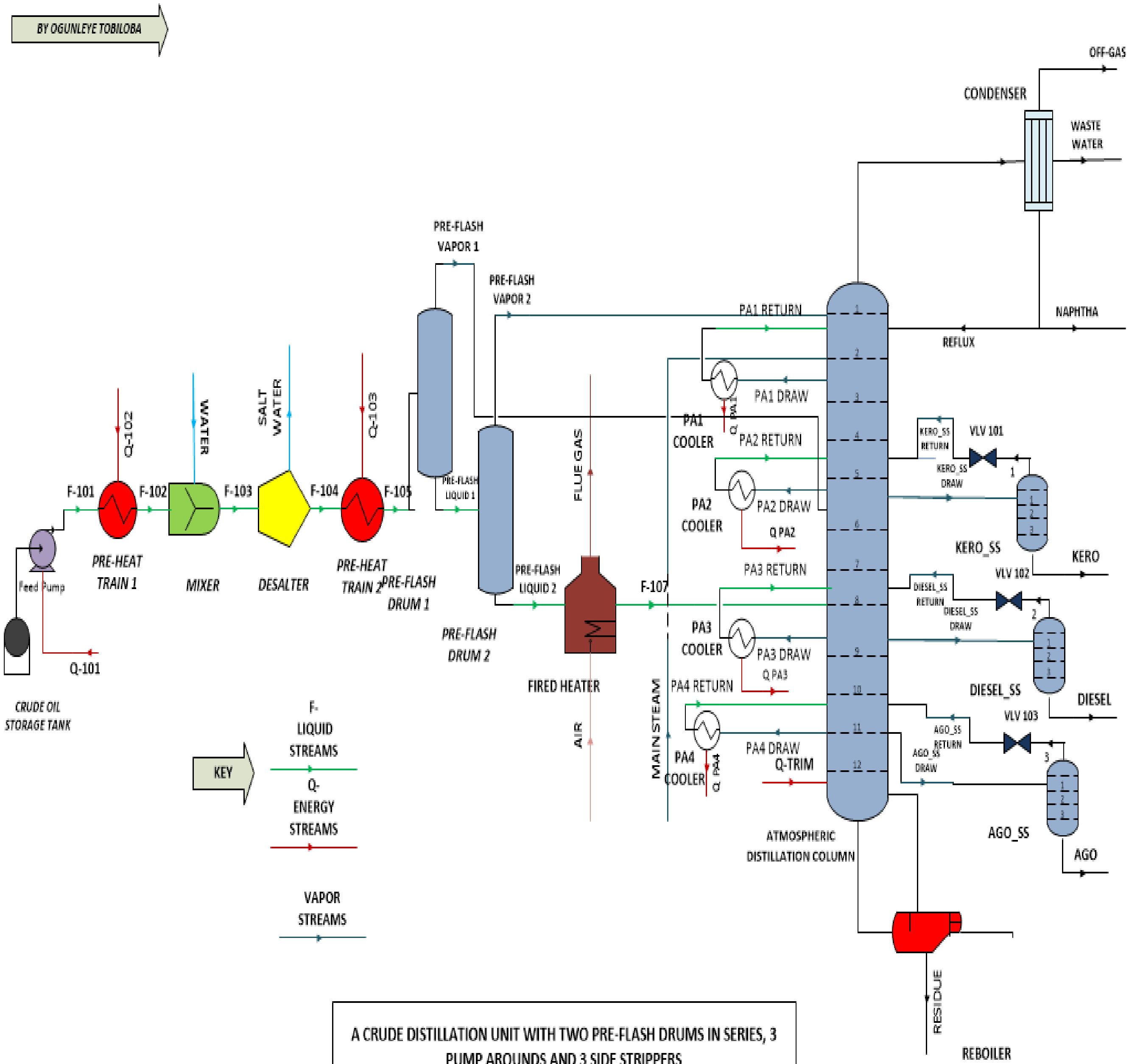


Figure 2. 6 PFD of a CDU with 2 pre-flash drums in series, 3 pump-around, and 3 side stripper

2.2.5 Modeling & Simulation of Modular Refineries

It was stated in the previous section that it is the responsibility of researchers and process engineers to design a new, more efficient, and less expensive refinery as well as to devise plans and procedures for the improved operation of an existing refinery. Changes in the refining units have a great impact on product yield and quality and it is therefore recommended to operate these units at optimal conditions from technical and economical points of view (i.e operating conditions such as temperature, pressure and flow rates of the units) that maximize their economic performance (increasing product yield), subject to their real physical restrictions and their design capabilities (Lopez *et al.*, 2009).

Process optimization and control cannot be physically done because it requires a large number of iterations that would be too expensive to perform, prone to human errors, and waste time hence the need for mathematical modeling & computer simulation of the system (Dauda, Megan and Gonzalo, 2017). A model is a mathematical representation of the physical and chemical phenomena taking place in a system or the relationship that describes what is happening in the system, while the activities leading to the construction of a model is referred to as modeling (Stephanopoulous, 1983). The two types of model available for the design of a crude distillation unit are the simple and rigorous models.

The shortcut (simple) distillation models adapt the Fenske–Underwood–Gilliland design equations for simple distillation columns by predicting the number of trays in each column section and operating conditions, such as reflux and re-boil ratios. The allowable configurations and accuracy of these models are inadequate (Optimization Based Design, 2017). The rigorous models apply mass and energy balances as well as equilibrium relations in every stage of the column. They provide more accurate predictions when compared to that of simple models. However, they are more difficult to handle due to the need to start the calculations from a very good initial guess in order to avoid convergence problems (Gadalla *et al.*, 2015).

Procedures for solving rigorous models are well established, and have been implemented in commercial process simulation software such as Aspen HYSYS, Aspen Plus, UNISIM, and PRO II. Such software allows designers to simulate complex distillation column flowsheets using iterative and sequential modular algorithms (Ledezma-martínez, Jobson and Smith, 2018). Over the years, process engineers have modeled and simulated various refineries for the purpose of optimization.

For instance,.(Handogo, 2012) simulated a modular refinery processing 30,000 bpd oil of Indonesian origin. The objective was to get the optimum temperature of the pre-flash column furnace that would influence its flash zone temperature. He observed that when intermediate product & gas oil molar flow rates and quality of heavy naphtha were kept constant, the optimum temperature was 573K with a profit of US$ 9.62 per m^3 compared to when the optimum temperature of 533K with profit of US$ 7.74 per m^3 when heavy naphtha & kerosene distillate molar flow rates and intermediate product quality were kept constant.

(Gadalla *et al.*, 2015) proposed a procedure to simulate an existing modular refinery processing 100,000bpd oil of Arabian origin. The objective was to increase energy efficiency, optimize operating conditions and reduce CO_2 emission from the refinery by either optimizing the existing pump-around and heat exchanger networks, installing a new pump-around or the installation of a pre-flash drum at a specific position. Among these three optimization techniques, the installation of a pre-flash drum was found to be the most effective.

(Kandasamy *et al.*, 2019) also simulated a simple crude distillation unit to process different types of crude oil. The aim of the simulation was to increase the Liquefied Petroleum Gas (LPG) yield from the CDU by optimizing the furnace temperature, number of stages, and steam to feed ratio. They assumed that there was no accumulation of mass in the CDU, no heat loss during the process, the effect of recycle stream was negligible, no unreacted feed present in the tower, and the heat capacity of all streams and column contents were considered constant. Results showed that the number of stages was the most significant factor that affected the LPG yield.

(Mamudu, Igwe and Okonkwo, 2019) simulated a crude distillation unit (CDU) that could process different crude oil assays of Nigerian origin. Simulations of these different crude assays were carried out on three different modular refinery configurations; simple, pre-flash, and pre-flash pump around reflux to determine which configuration would give the maximum product yield of each assay. They noticed that simple distillation & pre-flash scheme favors the production of diesel and residue oil for NDPR crude assay; gas oil and diesel oil for the Bonny light, qua-iboe, Bonny medium and brass crude assay while the pre-flash with pump-around scheme favors production of light naphtha for all crude assays; production of LPG for Bonny medium, qua-iboe and brass crude assay.

2.2.6 Optimization Studies of a Modular Refinery

It has been established that proper optimization studies require an accurate modeling & simulation of the plant and also efficient mathematical programming technique to achieve an optimal solution.

There are several strategies that can be used to perform mathematical optimization. In the past years, Linear Programming (LP) algorithms had been used to optimize simple modular refineries described with linear models (Lopez *et al.*, 2009). However, the introduction of pre-heating trains, pump-around, side strippers and pre-flash drums gave rise to non-linear variables that led to non-real solutions on the application of the LP technique (Barton *et al.*, 2015).

Rigorous models developed to properly represent the non-linear variables were linked to an optimizer (Basak 2002). Although this type of optimization began with a precise CDU representation, there were convergence problems and calculation time increased significantly for very complex models (Lopez *et al.*, 2009).

(Zhang and Zhu, 2000), optimized the operating temperature of a 100,000 bpd oil refinery using conventional linear programming (LP), site-level (master model) optimization, and process- level (sub model) optimization so as to maximize the overall annual profit which resulted in an increase of $17.5 million to $29.57 million, $20.87 million, and $29.89 million respectively. The increase in profit achieved from the LP algorithm was unrealistic as the model had significant simplifications. The integration of site- and process-level provides significant synergy for economic development.

(Zhang, Zhu and Towler, 2001) then furthered their work by proposing another strategy that applies the LP algorithm to include non-linear process models which resulted in about $1.9 million more profit when compared to the previous algorithms. This concept was implemented in some commercial software such as Aspen-PIMS, thus creating the evolution of such problems.

(Lopez *et al.*, 2009) proposed a Non-linear Programming (NLP) algorithm for an atmospheric, and vacuum distillation system with constant composition of crude feedstock using the meta-model approach to represent the non-linear phenomenon of distillation in order to maximize the system profit by finding the optimum operational conditions for the atmospheric tower, calculating products yields and their properties, temperatures and duties of exchangers responsible for crude pre-heating. There was an increment in profit of about $200 million dollars annually without modifications in installed infrastructure.

(Barton *et al.*, 2015) presented a mixed-integer nonlinear programming (MINLP) algorithm and combined it with 3 different advanced interval reduction (IR) techniques in global optimization software to maximize the profit by optimizing the amount of crude purchased and refinery operating conditions. Purchasing 40 Kilo-tons of crude 1, 200 Kilo-tons of crude 2, and 200 Kilo-tons of crude 3 resulted in a monthly profit of $31.8 million. The computational time reduced from 6544s when pure IR was used to 1261s when feasibility-based IR and optimality-based IR was used.

(Dauda, Megan and Gonzalo, 2017) applied rigorous tray-by-tray distillation column models to simulate a CDU. These models are combined with a genetic algorithm that optimizes the column design. The number of trays in each column section together with the operating conditions (including the feed inlet temperature, pump-around duties and temperature drops, stripping steam flow rates and reflux ratios) are selected to minimize the total annualized cost. Total annual cost reduced from $10.87 million to $8.47 million.

(Ledezma-martínez, Jobson and Smith, 2018) proposed a systematic design optimization approach for crude oil distillation systems with pre-flash unit, applying a rigorous simulation model and using pinch analysis to determine the minimum hot utility demand of the heat-integrated system. The code of genetic algorithm is solved using Matlab R2016a software after the system has been modeled in Aspen Hysys. The hot utility requirement was reduced from 58.3 MW to 46.6 MW with no pre-flash and 37.9 MW with pre-flash thereby saving costs.

(Ahmed and Khalaf, 2015), (Amin *et al.*, 2018), and (Ahmad *et al.*, 2018) developed other algorithms for optimization studies such as Taguchi method of cut-point temperature, hybridization of Taguchi method and genetic algorithms, artificial neural networks (ANN) and artificial neural networks based on Monte Carlo simulation (ANNBMC) amongst others.

2.2.7 Control of a Modular Refinery

The ultimate goal of the optimum configuration, modeling and simulation, and optimization studies of a modular refinery is to develop a control system that satisfies several requirements imposed on it by its designers and the general technical, economic, and social conditions in the presence of ever-changing external influences (disturbances). Among such requirements are safety, product specification, environmental regulations, operational constraints, economics and so on (Stephanopoulous, 1983). All the requirements listed above dictate the need for continuous monitoring of the operation of a chemical plant and external intervention (control) to guarantee the satisfaction of its operational objectives. This is accomplished through a rational arrangement of equipment (measuring devices, valves, controllers, and computers) and human intervention (plant designers and operators) which together constitute the control system (Serdar *et al.*, 2004).

The three basic needs a control system has to satisfy are suppressing the influence of external disturbances, ensuring the stability of a chemical process, and optimizing the performance of a chemical process. In the design of a control system, the control engineer is concerned about the;

i) Operational objective that the control system is set to achieve,

ii) Variables that should be measured in order to monitor the operational performance of the plant,

iii) Manipulated variables to be used to control a chemical process,

iv) Best control configuration for a given chemical process control situation (Single Input Single Output, SISO or Multiple Input Multiple Output, MIMO),

v) How information taken from measurements will be used to adjust the manipulated variables (Implementation by the controller).

(Gonçalves, Martins and Azevedo, 2010) developed a dynamic model with a suitable MIMO control configuration that represents an atmospheric distillation tower to evaluate the transient behavior of the unit for different operational situations. The controllers implemented the Proportional-Integral-Derivative (PID) control algorithm. The manipulated variables were the quality of the products (naphtha, kerosene & diesel), temperature (furnace & pump-around) and pressure (condenser). A step change was applied to each of the manipulated variable separately after 15 minutes of steady operations and the simulations were allowed to run until a new steady state conditions were reached.

(Bassem-Sayed *et al.*, 2016) also simulated an existing crude distillation unit with a MIMO control configuration to optimize the process. The controllers implemented the P, PI, and PID control algorithm. The manipulated variables were temperature (residue, pump-around, and feed), pressure (condenser), level (condenser and distillation column), and flow rate (crude oil, kerosene, and gas oils). A step change was applied to each manipulated variable and it was found that the

PID controller had a rapid response but was more difficult to tune when compared with the P and PI controllers.

(Serdar *et al.*, 2004) applied a model predictive control (MPC) algorithm to a crude distillation unit of an existing refinery which comprises of a base layer of PID control algorithm. All PID controllers were retuned before the MPC testing started. The manipulated variables selected were flow control (feed, furnace oil), pressure control (distillation column), and temperature control (distillation column). The disturbance variables selected were amount of total feed and inlet flow rates into the two furnaces. The controlled variables were temperature (desalter and furnace outlet), pressure (desalter), product quality (naphtha, diesel, kerosene), and operational constraints. Dynamic step changes were applied to the manipulated variables and the corresponding response (significant increase in product throughput and other economic objectives) validated the controller.

(Shigueo, Carlos and Odloak, 2015) also applied a model predictive control algorithm based on input targets and zone control to an existing crude distillation unit. The manipulated variables are flowrate (crude oil, stripping steam, pump-around, kerosene), temperature (pre-flash, distillation column). The controlled variables are flowrate (naphtha, top reflux of pre-flash & distillation column), quality (naphtha, diesel), and heat duty of furnace. The MPC algorithm was tuned by the real time optimization technique and the desired set points were determined for optimization.

CHAPTER THREE - METHODOLOGY

The modeling, simulation and control of a modular refinery would consist of the following phases:

Development of a transfer function model that describes the behavior of the modular refinery: Includes material and energy balances for every unit operation, equilibrium relations for all columns, and Laplace transforms to obtain the transfer function model in the s-domain.

Simulation of the transfer function model in Matlab-Simulink software so as to obtain the system response to step changes in input variables.

Design of a suitable proportional, integral and derivative (PID) controller to be tuned using Ziegler-Nicholas, Cohen-Coon, Integral Time Weighted Absolute Error (ITAE), and Internal Model Control (IMC) tuning methods.

3.1 Development of Transfer Function Model of a Modular Refinery Using Input / Output Method

3.1.1 Total Dynamic Mass Balance

[Rate of change of mass of fluid in system] = [Mass flow rate of fluid into system] - [Mass flow rate of fluid out of system]

$$\frac{d(\rho V)}{dt} = \Sigma_{\text{i: inlet}}\ \rho_i F_i - \Sigma_{\text{j: outlet}}\ \rho_j F_j \qquad (1.1.1)$$

3.1.2 Component 'A' Dynamic Mass Balance

$$\frac{d(Ca\,V)}{dt} = \Sigma_{\text{i: inlet}}\ C_{Ai} F_i - \Sigma_{\text{j: outlet}}\ C_{Aj} F_j \pm rV \qquad (1.2.1)$$

3.1.3 Total Dynamic Energy Balance

[Accumulation of total energy/Time] = [Input of total energy/Time] – [Output of total energy/Time]

$$\frac{d(\rho\, V\, Cp\, T)}{dt} = \rho\, F_i\, C_p\, T_i + Q + \rho\, F\, C_p\, T \qquad (1.3.1)$$

3.1.4 Pre-Heat Trains

The pre-heat trains will be modeled as a constant volume stirred heated tank

Total Dynamic Mass Balance:

$$\frac{d(\rho V)}{dt} = \rho\, F_i - \rho\, F \qquad (1.4.1)$$

$$\rho \frac{dV}{dt} = \rho\, (F_i - F)$$

Assumption: Constant Density

$$\frac{dV}{dt} = F_i - F$$

But V = Ah

$$\frac{d(Ah)}{dt} = F_i - F$$

$$A\frac{dh}{dt} = F_i - F \qquad (1.4.2)$$

Total Dynamic Energy Balance:

$$\frac{d}{dt}(\rho\ V\ C_p\ T) = \rho\ F_i\ C_p\ T_i - \rho\ F\ C_p\ T + Q \qquad (1.4.3)$$

Density ρ , Specific Heat Capacity C_p , and Cross-sectional area A are constant

$$\rho\ C_p\ A\ \frac{d}{dt}(hT) = \rho\ C_{pi}\ F_i\ T_i - \rho\ C_p\ F\ T + Q \qquad (1.4.4)$$

Dividing (1.4.4) through by $\rho\ C_p$, gives:

$$A\frac{d}{dt}(hT) = F_i\ T_i - F\ T + \frac{Q}{\rho\ Cp} \qquad (1.4.5)$$

$$\text{But } A\frac{d}{dt}(hT) = Ah\ \frac{dT}{dt} + AT\ \frac{dh}{dt} \qquad (1.4.6)$$

Substituting (1.4.2) & (1.4.6) into (1.4.5) gives:

$$Ah\ \frac{dT}{dt} + T(F_i - F) = F_i\ T_i - F\ T + \frac{Q}{\rho\ Cp} \qquad (1.4.7)$$

$$Ah\ \frac{dT}{dt} = F_i\ (T_i - T) + \frac{Q}{\rho\ Cp}$$

$$\frac{dT}{dt} = \frac{Fi\ Ti}{V} + \frac{F\ T}{V} + \frac{Q}{\rho\ Cp\ V}$$

For a constant volume heater, $F_i = F$;

If $\frac{V}{F} = \tau$, then;

$$\frac{dT}{dt} = \frac{1}{\tau}\ T_i - \frac{1}{\tau}\ T + \frac{Q}{\rho\ Cp\ V} \qquad (1.4.8)$$

At Steady-State, (1.4.8) becomes;

$$\frac{dTs}{dt} = \frac{1}{\tau}\ T_{is} - \frac{1}{\tau}\ T_s + \frac{Qs}{\rho\ Cp\ V} \qquad (1.4.9)$$

Subtracting (1.4.9) from (1.4.8) gives;

$$\frac{d}{dt}(T - T_s) = \frac{1}{\tau}(T_i - T_{is}) - \frac{1}{\tau}(T - T_s) + \frac{1}{\rho\, Cp\, V}(Q - Q_s) \qquad (1.4.10)$$

Introducing Deviation Variables;

$y = T - T_s$

$u = Q - Q_s$

$d = T_i - T_{is}$

$$\frac{dy}{dt} = \frac{1}{\tau} d - \frac{1}{\tau} y + \frac{1}{\rho\, Cp\, V} u \qquad (1.4.11)$$

Multiplying (1.4.11) through by τ gives;

$$\tau \frac{dy}{dt} + y = d + \frac{\tau}{\rho\, V\, Cp} u \qquad (1.4.12)$$

If $\frac{1}{\rho\, F\, Cp} = K$, then (1.4.12) becomes;

$$\tau \frac{dy}{dt} + y = d + K u \qquad (1.4.13)$$

Representing (1.4.13) in s-domain by taking its laplace transform gives;

$$y(s) = \frac{k}{\tau s+1} u(s) + \frac{1}{\tau s+1} d(s) \qquad (1.4.14)$$

3.1.5 Furnace

Total Dynamic Mass Balance;

$$\frac{d}{dt}(\rho\, V) = \rho\, F_i - \rho\, F + Q_F \qquad (1.8.1)$$

Assuming Constant Density,

$\rho \frac{dV}{dt} = \rho F_i - \rho F + Q_F$

$\frac{dV}{dt} = F_i - F + Q_F$ (1.8.2)

Total Dynamic Energy Balance;

$\frac{d}{dt}(\rho V C_p T) = \rho F_i C_p T_i - \rho F C_p T + Q_F$ (1.8.3)

$\rho V C_p \frac{dT}{dt} = \rho F_i C_p T_i - \rho F C_p T + Q_F$

$\frac{dT}{dt} = (F_i T_i / V) - (F T / V) + (Q_F / \rho VC_p)$

Let $\tau = V / F$ and $F_i = F$

$\frac{dT}{dt} = \frac{1}{\tau} T_i - \frac{1}{\tau} T + (Q_F / \rho V C_P)$ (1.8.4)

At steady state, (1.8.4) becomes;

$\frac{dT,s}{dt} = \frac{1}{\tau} T_{i,s} - \frac{1}{\tau} T_s + (Q_{F,s} / \rho V C_P)$ (1.8.5)

Subtracting (1.8.5) from (1.8.4) gives;

$\frac{d(T - T,s)}{dt} = \frac{1}{\tau} (T_i - T_{i,s}) - \frac{1}{\tau} (T - T_s) + (Q_F - Q_{F,s}) / \rho V C_P)$ (1.8.6)

Introducing Deviation Variables;

$y = T - T_s$

$u = Q_F - Q_{F,s}$

$d = T_i - T_{i,s}$

Therefore, (1.8.6) becomes;

$\frac{dy}{dt} = \frac{1}{\tau} d - \frac{1}{\tau} y + \frac{1}{\rho V Cp} u$ (1.8.7)

Rearranging (1.8.7) gives;

$\tau \frac{dy}{dt} + y = K.u + d$ (1.8.8)

Where $K = \frac{1}{\rho F Cp}$

Representing (1.8.8) in s-domain by Laplace Transforms gives;

$y(s) = \frac{K}{\tau s+1} u(s) + \frac{1}{\tau s+1} d(s)$ (1.8.9)

$$y\,(s) = \frac{K}{\tau s + 1} u(s) + \frac{1}{\tau s + 1}$$

3.1.9 Distillation Column

Assumptions;

1. The liquid on each tray is perfectly mixed & of uniform composition X_n
2. There are negligible heat losses from the column to the atmosphere.
3. The vapor holdup h_n on each tray is negligible i.e $V = V_1 = V_2 = V_n$.
4. Efficient trays
5. Vapor and liquid are in thermal equilibrium ($T_V = T_L$) but not in phase equilibrium, so Murphree efficiency will be used to describe its departure from equilibrium (Non-equimolal overflow).

6. Dynamics of the condenser & reboiler will be neglected. This is because in commercial-scale columns, the dynamic response of these heat exchangers is usually much faster than the response of the column.
7. Liquid hydraulics are calculated from Francis Weirs formula.
8. Pressure is constant and known on each tray. It varies linearly up the column

Total mass balance, component mass balance and total energy balance will be written separately for the;

1. Condenser / Receiver
2. Top tray 'N'
3. Arbitrary tray 'n'
4. Feed tray 'f'
5. The first tray in the stripping section (Tray 1)
6. The reboiler and column base.

3.1.9.1 CONDENSER

Total Dynamic mass balance;

$\frac{d}{dt}H_D = V - R - D$ (1.9.1.1)

Component A Dynamic mass balance;

$\frac{d}{dt}H_D X_D = V\, y_n - (R + D)\, X_D$ (1.9.1.2)

But $\frac{d}{dt}H_D X_D = X_D \frac{d}{dt}H_D + H_D \frac{d}{dt}X_D$ (1.9.1.3)

Substituting (1.9.1.3) and (1.9.1.1) into (1.9.1.2) gives;

$$X_D (V - R - D) + H_D \frac{d}{dt} X_D = V\, y_n - (R + D)\, X_D \qquad (1.9.1.4)$$

Rearranging (1.9.1.4) gives;

$$\frac{d}{dt} X_D = V\, y_n / H_D - V\, X_D / H_D \qquad (1.9.1.5)$$

Let $\tau = H_D / V$, then

$$\frac{d}{dt} X_D = -\frac{1}{\tau} X_D + \frac{1}{\tau} y_n \qquad (1.9.1.6)$$

At steady state, (1.9.1.6) becomes;

$$\frac{d}{dt} X_{D,s} = -\frac{1}{\tau} X_{D,s} + \frac{1}{\tau} y_{n,s} \qquad (1.9.1.7)$$

Subtracting (1.9.1.7) from (1.9.1.6) gives;

$$\frac{d}{dt} (X_D - X_{D,s}) = -\frac{1}{\tau} (X_D - X_{D,s}) + \frac{1}{\tau} (y_n - y_{n,s}) \qquad (1.9.1.8)$$

Introducing deviation variables;

$y = X_D - X_{D,s}$

$u = (y_n - y_{n,s})$

$$\frac{dy}{dt} = -\frac{1}{\tau} y + \frac{1}{\tau} u$$

$$\tau \frac{dy}{dt} + y = K.u \qquad (1.9.1.9)$$

Where; $K = 1$

Representing (1.9.1.9) in s-domain by taking its Laplace transform gives;

$y(s) = \frac{K}{\tau s+1} u(s)$

3.1.9.2 TOP TRAY 'N'

Total Dynamic Mass Balance;

$$\frac{d}{dt}H_N = V_{n-1} - L_n - V_n + R \qquad (1.9.2.1)$$

From assumption (3), $V_{n-1} = V_n$, (1.9.2.1) becomes;

$$\frac{d}{dt}H_N = R - L_n \qquad (1.9.2.2)$$

Component A Mass Balance ;

$$\frac{d}{dt}H_N X_N = V y_{n-1} - L_n X_n - V_n y_n + R X_D \qquad (1.9.2.3)$$

But

$$\frac{d}{dt}H_N X_N = H_N \frac{d}{dt} X_N + X_N \frac{d}{dt} H_N \qquad (1.9.2.4)$$

Substituting (1.9.2.2) and (1.9.2.4) into (1.9.2.3) gives;

$$X_N (R - L_N) + H_N \frac{d}{dt} X_N = V y_{n-1} - L_n X_n - V_n y_n + R X_D \qquad (1.9.2.5)$$

Rearranging (1.9.2.5) gives;

$$\frac{d}{dt}X_N = V / H_N (y_{n-1} - y_n) + R X_D / H_N - R X_N / H_N \qquad (1.9.2.6)$$

At steady state, (1.9.2.6) becomes;

$$\frac{d}{dt}X_{N,s} = V_s / H_N (y_{n-1} - y_n) + (R X_{D,s}) / H_N - (R X_{N,s} / H_N) \qquad (1.9.2.7)$$

Subtracting (1.9.2.7) from (1.9.2.6) gives;

$\frac{d}{dt}(X_N - X_{N,s}) = [(V - V_s)(y_{n-1} - y_n)] / H_N + (X_D - X_{D,s}) R / H_N - (X_N - X_{N,s}) R / H_N$ (1.9.2.8)

Introducing Deviation Variables;

$y = X_N - X_{N,s}$

$u = V - V_s$

$d = X_D - X_{D,s}$

$\tau = H_N / R$

$\frac{dy}{dt} = -\frac{1}{\tau} y + \frac{1}{\tau} d + [(y_{n-1} - y_n) / H_N]\, u$

$\tau \frac{dy}{dt} + y = K.u + d$ (1.9.2.9)

Where $K = (y_{n-1} - y_n) / R$

Representing (1.9.2.9) in s-domain by taking its Laplace transform gives;

$$y(s) = \frac{\boldsymbol{K}}{\boldsymbol{\tau s + 1}} \boldsymbol{u}(\boldsymbol{s}) + \frac{\mathbf{1}}{\boldsymbol{\tau s + 1}} \boldsymbol{d}(\boldsymbol{s})$$

3.1.9.3 ARBITRARY TRAY 'N'

Total Dynamic mass balance;

$\frac{d}{dt} H_n = L_{n+1} + V_{n-1} + V_{PF} - V_n - L_n - S_n$ (1.9.3.1)

From assumption 3, $V_{n-1} = V_n$, (1.9.3.1) then becomes;

$$\frac{d}{dt} H_n = L_{n+1} + V_{PF} – L_n – S_n \qquad (1.9.3.2)$$

Component ‘A’ balance;

$$\frac{d}{dt} H_n X_n = L_{n+1} X_{n+1} + V_{n-1} y_{n-1} + V_{PF} y_{PF} – V_n y_n – L_n X_n – S_n X_n \qquad (1.9.3.3)$$

But

$$\frac{d}{dt} H_N X_N = H_N \frac{d}{dt} X_N + X_N \frac{d}{dt} H_N \qquad (1.9.3.4)$$

Substituting (1.9.3.2) and (1.9.3.4) into (1.9.3.3) gives;

$$H_N \frac{d}{dt} X_N + X_N (L_{n+1} + V_{PF} – L_n – S_n) = L_{n+1} X_{n+1} + V_{n-1} y_{n-1} + V_{PF} y_{PF} – V_n y_n – L_n X_n – S_n X_n \qquad (1.9.3.5)$$

Rearranging (1.9.3.5) gives;

$$\frac{d}{dt} X_n = -X_n (L_{n+1} + V_{PF}) / H_N + (V\, y_{n-1} / H_N) + (L_{n+1} X_{n+1} / H_N) + (V_{PF}\, y_{PF} / H_N) \qquad (1.9.3.6)$$

If $V_{PF} = 0$, i.e there is no vapor from Pre-flash drum entering the arbitrary tray n, (1.9.3.6) becomes;

$$\frac{d}{dt} X_n = -X_n (L_{n+1}) / H_N + (V\, y_{n-1} / H_N) + (X_{n+1} L_{n+1} / H_N) \qquad (1.9.3.7)$$

At steady state, (1.9.3.7) becomes;

$$\frac{d}{dt} X_{nS} = -X_{nS} (L_{n+1}) / H_N + (V_S\, y_{n-1} / H_N) + (X_{n+1,S} L_{n+1} / H_N) \qquad (1.9.3.8)$$

Subtracting (1.9.3.8) from (1.9.3.7) gives ;

$$\frac{d}{dt}(X_n - X_{nS}) = -(X_n - X_{nS}) (L_{n+1}) / H_N + (V - V_S) y_{n-1} / H_N) + (X_{n+1} - X_{n+1,S}) L_{n+1} / H_N) \qquad (1.9.3.9)$$

Introducing Deviation Variables;

$y = X_n – X_{nS}$

$u = V – V_S$

$d = X_{n+1} - X_{n+1,S}$

$\tau = H_N / L_{n+1}$

$$\frac{dy}{dt} = -\frac{1}{\tau} y + \frac{1}{\tau} d + (y_{n-1} / H_N) u \qquad (1.9.3.10)$$

Rearranging (1.9.3.10) gives;

$$\tau \frac{dy}{dt} + y = K.u + d \qquad (1.9.3.11)$$

Where $K = (y_{n-1} / L_{n+1})$

Representing (1.9.3.11) in s-domain by taking its Laplace transform gives;

$$y(s) = \frac{K}{\tau s+1} u(s) + \frac{1}{\tau s +1} d(s)$$

3.1.9.4 FEED TRAY 'N'

Total Dynamic Mass Balance

$$\frac{d}{dt}H_f = L_{f+1} + V_{f-1} + F + V_f – L_f \qquad (1.9.4.1)$$

From assumption 3, $V_{f-1} = V_f = V$, (1.9.4.1) becomes;

$$\frac{d}{dt}H_f = L_{f+1} + F – L_f \qquad (1.9.4.2)$$

Component A Mass Balance;

$$\frac{d}{dt}(H_f X_f) = L_{f+1} X_{f+1} + V y_{f-1} + F X_f + V y_f – L_f X_f \qquad (1.9.4.3)$$

But

$$\frac{d}{dt}H_f X_f = H_f \frac{d}{dt} X_f + X_f \frac{d}{dt} H_f \qquad (1.9.4.4)$$

Substituting (1.9.4.2) and (1.9.4.4) into (1.9.4.3) gives;

$$H_f \frac{d}{dt} X_f + X_f (L_{f+1} + F – L_f) = L_{f+1} X_{f+1} + V y_{f-1} + F X_f + V y_f – L_f X_f \qquad (1.9.4.5)$$

Rearranging (1.9.4.5) gives;

$$\frac{d}{dt}X_f = - X_f (L_{f+1} / H_f) + X_{f+1} (L_{f+1} / H_f) + V (y_{f-1} – y_f) /H_f \qquad (1.9.4.6)$$

At steady state, (1.9.4.6) becomes;

$$\frac{d}{dt}X_{f,S} = - X_{f,S} (L_{f+1} / H_f) + X_{f+1,S} (L_{f+1} / H_f) + V_S (y_{f-1} – y_f) /H_f \qquad (1.9.4.7)$$

Subtracting (1.9.4.7) from (1.9.4.6) gives;

$$\frac{d}{dt}(X_f - X_{f,S}) = - (X_f - X_{f,S}) (L_{f+1} / H_f) + (X_{f+1} - X_{f+1,S}) (L_{f+1} / H_f) + (V - V_S) (y_{f-1} – y_f) /H_f \quad (1.9.4.8)$$

Introducing Deviation Variables;

$y= X_f - X_{f,S}$

$u= V - V_S$

$d= X_{f+1} - X_{f+1,S}$

$\tau = H_f / L_{f+1}$

$$\frac{dy}{dt} = - \frac{1}{\tau} y + \frac{1}{\tau} d + (y_{f-1} – y_f / H_f) u \qquad (1.9.4.9)$$

Rearranging (1.9.4.9) gives;

$$\tau \frac{dy}{dt} + y = K.u + d \qquad (1.9.4.10)$$

Where K= $y_{f-1} - y_f / L_{f+1}$

Representing (1.9.4.10) in s-domain by taking its Laplace transform gives;

$$y(s) = \frac{K}{\tau s+1} u(s) + \frac{1}{\tau s+1} d(s)$$

3.2.9.5 THE FIRST TRAY IN THE STRIPPING SECTION (TRAY 1)

Total Dynamic Mass Balance;

$$\frac{d}{dt}H_1 = L_2 + V_B - V_1 - L_1 \qquad (1.9.5.1)$$

From assumption 3, $V_B = V_1$, so (1.9.5.1) becomes;

$$\frac{d}{dt}H_1 = L_2 - L_1 \qquad (1.9.5.2)$$

Component A Dynamic Mass Balance;

$$\frac{d}{dt}(H_1X_1) = L_2 X_2 + V y_B - V y_1 - L_1 X_2 \qquad (1.9.5.3)$$

But

$$\frac{d}{dt}(H_1X_1) = X_1 \frac{d}{dt}H_1 + H_1 \frac{d}{dt}X_1 \qquad (1.9.5.4)$$

Substituting (1.9.5.2) and (1.9.5.4) into (1.9.5.3) gives;

$$H_1 \frac{d}{dt}X_1 + X_1 (\frac{d}{dt}H_1 = L_2 - L_1) = L_2 X_2 + V y_B - V y_1 - L_1 X_2 \qquad (1.9.5.5)$$

Rearranging (1.9.5.5) gives;

$$\frac{d}{dt}X_1 = -X_1\,(L_2 / H_1) + X_2\,(L_2 / H_1) + V\,(y_B - y_1) / H_1 \qquad (1.9.5.6)$$

At steady state, (1.9.5.6) becomes;

$$\frac{d}{dt}X_{1,S} = -X_{1,S}\,(L_2 / H_1) + X_{2,S}\,(L_2 / H_1) + V_S\,(y_B - y_1) / H_1 \qquad (1.9.5.7)$$

Subtracting (1.9.5.7) from (1.9.5.6) gives;

$$\frac{d}{dt}\,(X_1 - X_{1,S}) = -(X_1 - X_{1,S})\,(L_2 / H_1) + (X_2 - X_{2,S})\,(L_2 / H_1) + (V - V_S)\,(y_B - y_1) / H_1 \qquad (1.9.5.8)$$

Introducing Deviation Variables;

$y = X_1 - X_{1,S}$

$u = V - V_S$

$d = X_2 - X_{2,S}$

$\tau = H_1 / L_2$

$$\frac{dy}{dt} = -\frac{1}{\tau}\,y + \frac{1}{\tau}\,d + (\,y_B - y_1 / H_1\,)\,u \qquad (1.9.5.9)$$

Rearranging (1.9.5.9) gives;

$$\tau\,\frac{dy}{dt} + y = K.u + d \qquad (1.9.5.10)$$

Where $K = y_B - y_1 / L_2$

Representing (1.9.4.10) in s-domain by taking its Laplace transform gives;

$$y(s) = \frac{K}{\tau s+1}\,u(s) + \frac{1}{\tau s+1}\,d(s)$$

3.2.9.6 REBOILER AND COLUMN BASE

Total Dynamic Mass Balance;

$$\frac{d}{dt}H_B = L_1 - V - B \qquad (1.9.6.1)$$

Component A Dynamic Mass Balance;

$$\frac{d}{dt}H_B X_B = L_1 X_1 - V y_B - B X_B \qquad (1.9.6.2)$$

But

$$\frac{d}{dt}H_B X_B = H_B \frac{d}{dt}X_B + X_B \frac{d}{dt}H_B \qquad (1.9.6.3)$$

Substituting (1.9.6.1) and (1.9.6.3) into (1.9.6.2) gives;

$$H_B \frac{d}{dt}X_B + X_B (L_1 - V - B) = L_1 X_1 - V y_B - B X_B \qquad (1.9.6.4)$$

Rearranging (1.9.6.4) gives;

$$\frac{d}{dt}X_B = -X_B (L_1 - V) / H_B + X_1 (L_1 - V_0) / H_B + V y_B / H_B \qquad (1.9.6.5)$$

$V_0 = 0$, since there is no substantial vapor present in a liquid stream

At steady state, (1.9.6.5) becomes;

$$\frac{d}{dt}X_{B,S} = -X_{B,S} (L_1 - V) / H_B + X_{1,S} (L_1 - V_0) / H_B + V_S y_B / H_B \qquad (1.9.6.6)$$

Subtracting (1.9.6.6) from (1.9.6.5) gives;

$$\frac{d}{dt}(X_B - X_{B,S}) = -(X_B - X_{B,S}) (L_1 - V) / H_B + (X_1 - X_{1,S}) (L_1 - V_0) / H_B + (V - V_S) y_B / H_B \qquad (1.9.6.7)$$

Introducing Deviation Variables;

$y = X_B - X_{B,S}$

$u = V - V_S$

$d = X_1 - X_{1,S}$

$\tau = H_B / L_1 - V$

$\frac{dy}{dt} = - \frac{1}{\tau} y + \frac{1}{\tau} d + (y_B / H_B) u$ (1.9.6.8)

Rearranging (1.9.5.9) gives;

$\tau \frac{dy}{dt} + y = K.u + d$ (1.9.6.9)

Where $K = y_B / L_1 - V$

Representing (1.9.4.10) in s-domain by taking its Laplace transform gives;

$y(s) = \frac{K}{\tau s+1} u(s) + \frac{1}{\tau s +1} d(s)$

3.2 Data Generation for Crude Distillation Unit

Specific Heat Capacity Of Liquid Crude

Using the Fortsch et al correlation for calculating to calculate the specific heat of petroleum oils. (UNITED STATES DEPARTMENT OF COMMERCE, 1929)

$C_P = \frac{1}{\sqrt{d}} (0.388 + 0.00045t)$

Where C_P = Specific Heat Capacity of Petroleum Oil

d = Specific Gravity at 60°F

t = Temperature of crude in °F

Specific Gravity of Bonny Light Crude at 60°F = 0.8501901

Specific Heat Capacity of Liquid Outlet Petroleum Oil, C_{PL}

Series of Pre-heat Trains (T= 400 °F)

C_{PL} = 0.6160136 Btu/lbmol °F

Furnace (T= 630°F)

C_{PF} = 0.728263 Btu/lbmol °F

SPECIFIC GRAVITY OF CRUDE, S_G

Specific Gravity of Crude at 60°F, $S_{G60°F}$ = 0.8501901

$S_{GT} = S_{G60°F} - [3.31\times10^{-4} \times (T°F - 60)]$

Where T= Exit Temperature of Stream

Series of Pre-heat Trains (T = 400°F)

S_{GT} = 0.784789

Furnace (T = 630 °F)

S_{GT} = 0.6615201

Calculation of τ and K Values

1) Series of Pre-heat Trains

$\tau = \frac{V}{F}$

1 KJ/Kmol = 4.186 Btu/lbmol °F (Source: Perry's Chemical Engineering Handbook)
Volume of Pre-heater, V = 2m^3
Crude Feed into Pre-heater, F= 7.87037 kg/s (5000 BPD)

Therefore $\tau = \frac{2}{7.87037} = 0.2541176$

$$K = \frac{1}{\rho\, F\, C_p}$$

ρ = 0.6615201
C_{PF} = 0.616036 Btu/lbmol °F = 2.57863 KJ/Kmol

Therefore K= 0.06679

$$y(s)\frac{0.06679}{0.2541176s + 1}\, u\,(s)$$

2) Furnace
$\tau = \frac{V}{F}$
Area, A = 2.25m^2
Height, h = 1m
Volume, V = 2.25m^3
F = 7.87037 Kg/s
Therefore,
τ = 0.28588

$$K = \frac{1}{\rho\, F\, C_p}$$

ρ = 0.6615201

C_{PF} = 0.72863 Btu/lbmol °F = 3.04851 KJ/Kmol

Therefore,

K = 0.063

$$y(s) = \frac{0.063}{0.28588s + 1}\, u\,(s)$$

3) Distillation Column

Simulation of the distillation column was performed in Aspen Hysys to obtain compositions of the liquid and vapor streams necessary to define the transfer functions for each tray in the column. The steady state concentration X_n and Y_n necessary to define the transfer functions for each tray in the column were gotten from (Minh and Pumwa, 2012) and the column was simulated according to its specifications.

This paper presents control of level and temperature in the diesel chimney tray which is the choice product in Walter-Smith modular refinery. The modular refinery is made up of a 24-tray distillation column, series of pre-heaters which are fueled by side-products to increase heat re-integration, and a main heater / furnace. The feed enters at a temperature range of 625 °F – 630 °F while the top and bottom operating pressure are 5 bar and 7 bar respectively.

Diesel is drawn from tray 16 at a temperature of 520 °F and the transfer function for this tray will be calculated following the procedures in (Minh and Pumwa, 2012).

For Tray 16 in the distillation column;

$\tau = H_N / L_{n+1}$ and $K = (y_{n-1} / L_{n+1})$.

$H_N = 0.95\ \pi\ h_T\ D^{2} \times \frac{1}{4} \times \frac{d}{M.w}$

h_{16} = Average depth of pure liquid in Tray

$M.w_{16}$ = Molecular Weight of liquid held up on Tray

d_{16} = mean density of liquid held up (Diesel) on Tray

D = Diameter of the column = 1.558m

h_{16} = wet area / weir length = 1.678 m^2 / 1.2m = 1.398333 m

d = 690.20 Kg/m^3

M.w = 216.1 Kg/Kmole

Therefore H_N = 8.088710 Kmole

Molar Flowrates of Vapor & Liquid are constant,

Boil-up rate = 0.4048

Vapor rate = 0.4048 + 0.04 = 0.4448 (Addition of 4% is meant for tolerance of unexpected heavy components)

L_{17} = 100 – 44.48 = 55.52%

From literature, y_{17} = 0.8666

Therefore, $\tau = \frac{8.088710}{4.36962} = 1.8511203$

$K = \frac{0.8666}{4.36962} = 0.198186$

$$y(s) = \frac{0.198186}{1.8511203s+1} u(s) + \frac{1}{1.8511203s+1} d(s)$$

To obtain the process reaction curve for a diesel chimney tray, the controller is disconnected from the controlled process as shown in Figure 3.1 and a step change of magnitude A is implemented in the input to the controlled process via the final control element.

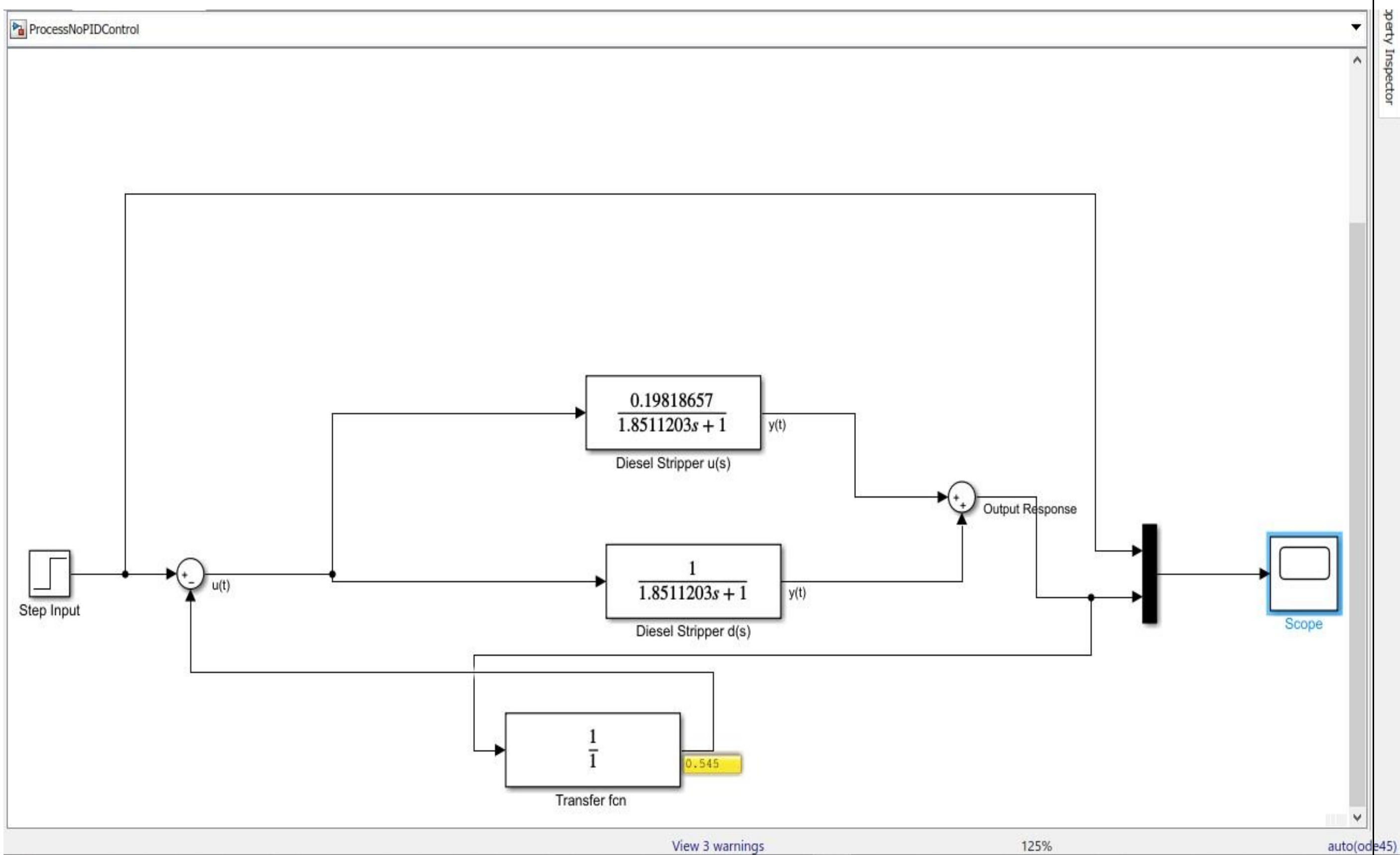


Figure 3. 1 Matlab-Simulink Simulation of the Modular Refinery without control

The resulting transient response or measured output gives the process reaction curve as shown in Figure 3.2. The next step is to estimate the characterizing parameters of the K, τ , and t_d to be used in tuning the controllers. This is achieved as shown in Figure 3.3 where a tangent line is drawn at the inflexion point on the process reaction curve Where this line intersects the time axis gives an estimate of the effective time delay.

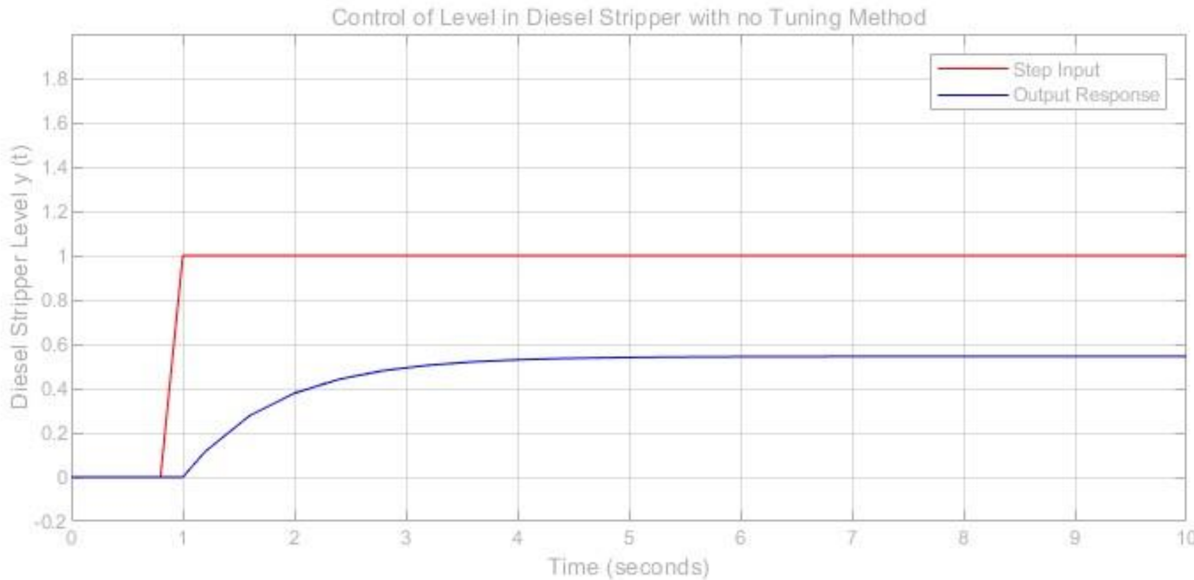


Figure 3. 2 Output Response of Level in Diesel Chimney Tray Without PID Control

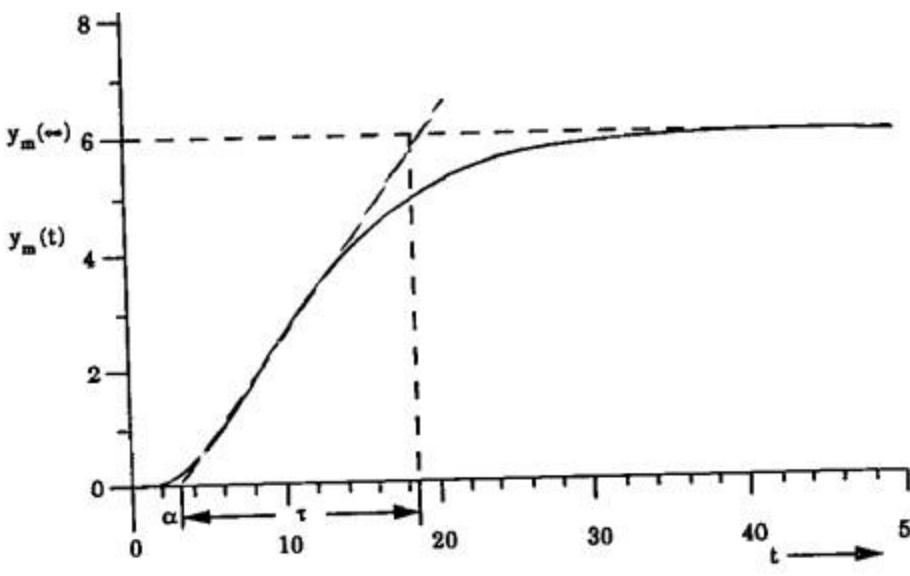


Figure 3. 3 Graphical Method of Estimating Characterization Parameters of a Controller

$$K = \frac{y_m}{A}$$

From Figure 3.2, Effective Gain, $K = \frac{0.55}{1} = 0.55$

Time Constant, $\tau = 2.5 - 1 = 1.5$

Time delay, $t_d = 1$

These values of K, τ, and t_d will be used in calculating K_C, τ_i, and τ_D values of four PID controller tuning methods : Ziegler-Nicholas (ZN), Cohen-Coon (CC), Integral Time Weighted Absolute Error (ITAE), and Internal Model Control (IMC) .

Ziegler Nicholas PID Controller Tuning Method

$$K_C = \frac{1.2}{K}\left(\frac{\tau}{t_d}\right) = 3.2727$$

$$\tau_I = 2 \times t_d = 2 \times 1 = 2$$

$$\tau_D = 0.5 \times t_d = 0.5 \times 1 = 0.5$$

Cohen-Coon PID Controller Tuning Method

$$K_C = \frac{1}{K}\left(\frac{\tau}{t_d}\right)\left(\frac{4}{3} + \frac{1}{4}\left({}^{t_d}/_{\tau}\right)\right) = 4.0909$$

$$\tau_I = t_d\left(\frac{32 + 6\left({}^{t_d}/_{\tau}\right)}{13 + 8\left({}^{t_d}/_{\tau}\right)}\right) = 1.9636$$

$$\tau_D = t_d\left(\frac{4}{11 + 2\left({}^{t_d}/_{\tau}\right)}\right) = 0.3243$$

ITAE PID Controller Tuning Method; Set Point

$$K_C = \frac{0.965}{K}\left(\frac{\tau}{t_d}\right)^{0.855} = 2.48154$$

$$\tau_I = \frac{\tau}{0.796 - 0.147\ \left(\frac{t_d}{\tau}\right)} = 2.14899$$

$$\tau_D = 0.308\ \tau\left(\frac{t_d}{\tau}\right)^{0.929} = 0.31699$$

IMC PID Controller Tuning Method

$\lambda > 0.2\ \tau$ and $\frac{\lambda}{t_d} > 0.25$

Therefore, let λ = 0.35

$$K_C = \frac{2\tau + t_d}{2K\ (\lambda + t_d)} = 2.6936$$

$$\tau_I = \tau + \frac{t_d}{2} = 2$$

$$\tau_D = \frac{\tau\ t_d}{2\tau + t_d} = 0.375$$

These four tuning methods were used in tuning the process to see which gives the desired dynamic response.

3.3 Control Scheme for a Modular Refinery

Control is necessary for the crude distillation unit in a refinery operation. The main objectives of applying control on processes are: maintaining safety, production specifications compliance with environment regulations, operational condition, and economic consideration

The control system developed for the modular refinery are flow, pressure, temperature, and level controls. The final control element for each control system and its output are listed below. Table 3.1 shows the controlled variables, manipulated variables and final control element of the modular refinery while Figure 3.4 shows the control scheme of the modular refinery. This paper focuses on the control of level in the diesel chimney tray as diesel is the most important product that comes out of the modular topping refinery.

Table 3. 1 Process Variables and Their Control Valves In A Modular Refinery

	Controlled Variables	Manipulated Variables	Control Valve
1	Distillation Column (Rectifying Section) Temperature	Reflux flow rate	TC-3
2	Distillation Column (Stripping Section) Temperature	Reboiler steam flow rate	TC-7
3	Distillation Column Feed Temperature	Fuel flowrate (Air to fuel ratio)	TC-2
4	Series of Pre-Heater Trains Temperature	Steam flow rate	TC-1
5	Condenser Pressure	Off-gas flow rate	PC-1
6	Distillation Column Level	Residue stream flow rate	LC-6
7	Condenser Level	Top product flow rate	LC-1
8	Side Strippers Level	Side products flow rate	LC-3, LC-4 & LC-5
9	Reflux Drum Level	Naphtha stream flow rate	LC-2
10	Distillation Column Feed Flow	Feed flow rate	FC-2

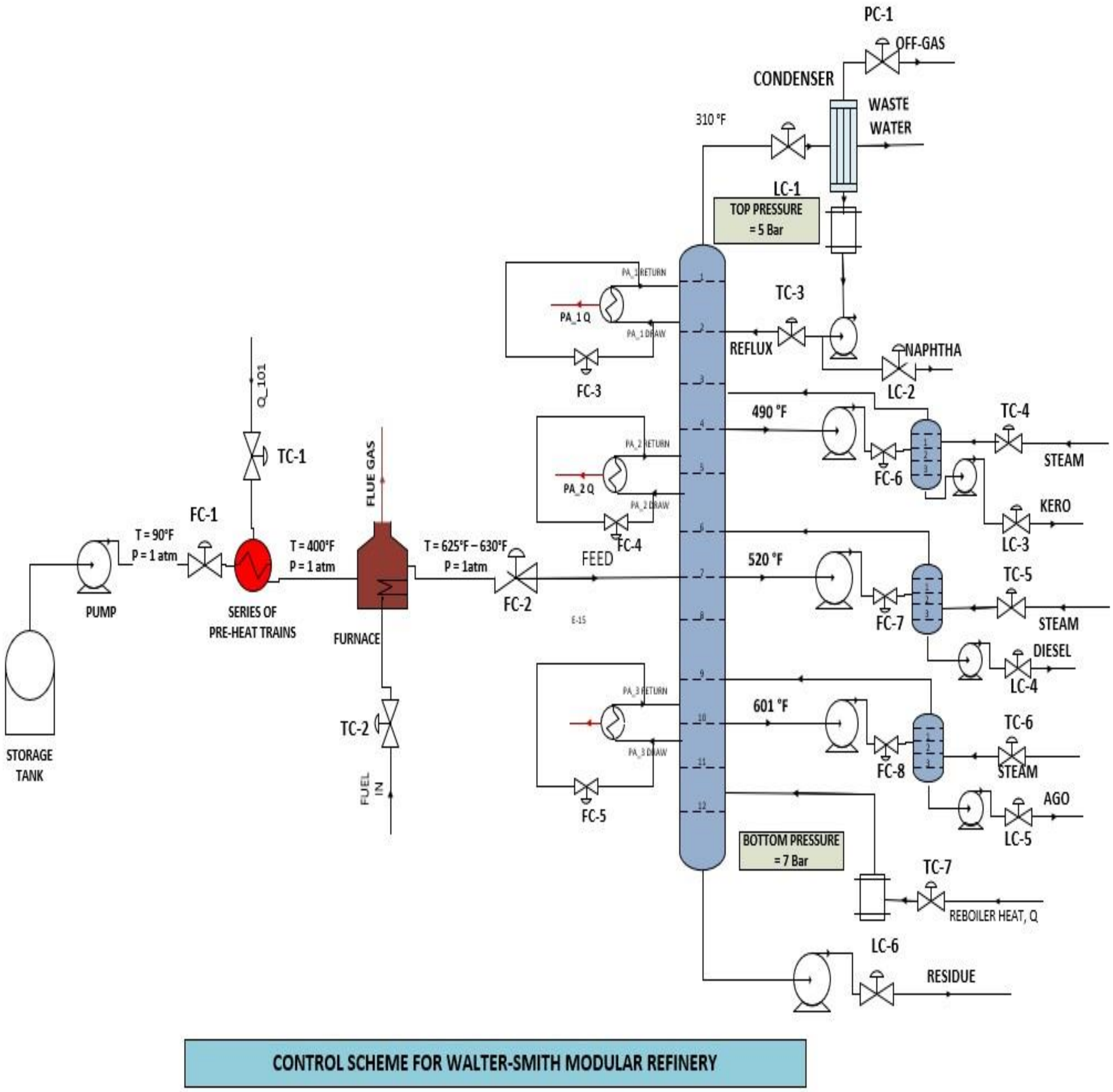


Figure 3. 4 Control Scheme for a Modular Refinery.

CHAPTER FOUR - RESULTS AND DISCUSSION

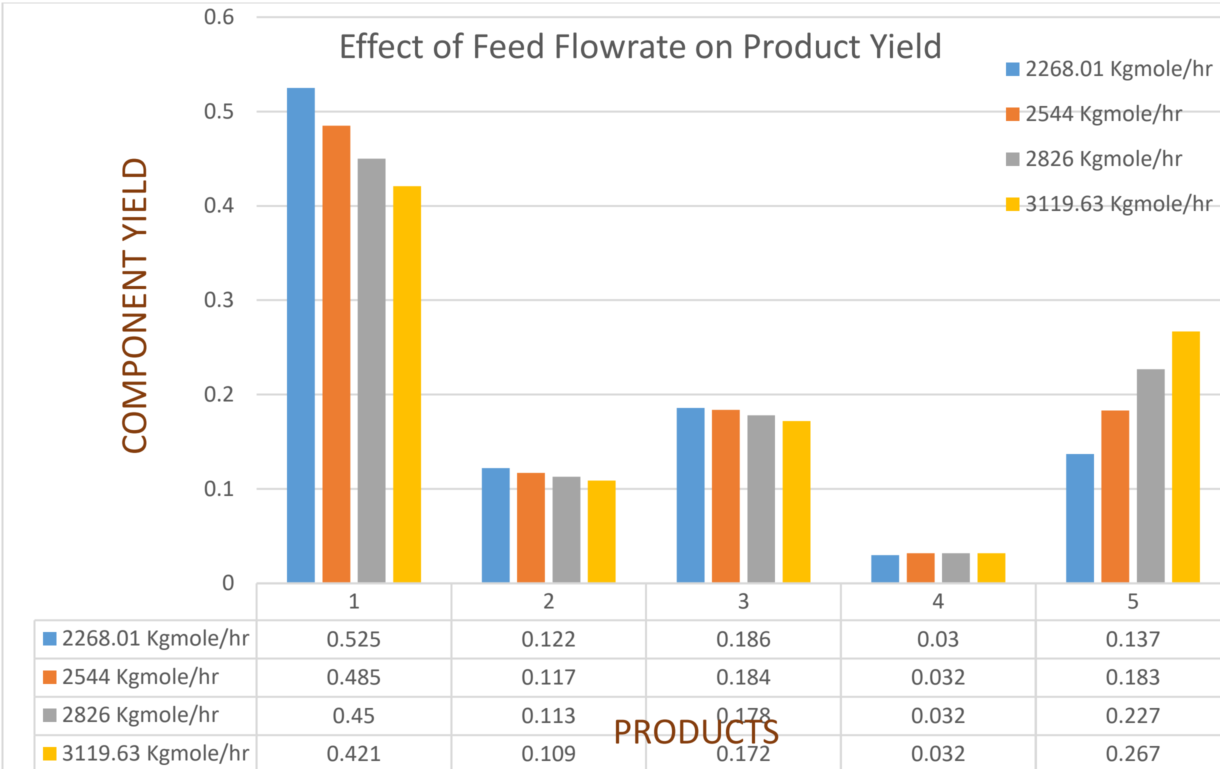
4.1 Effect of Feed Flow Rate on Product Yield

The feed flow rate is the rate at which feed mixture (liquid and vapor) from furnace and pre-flash drum respectively is pumped into the column. The feed flow rate can affect the efficiency of the column, quantity and quality of products / fractions.

Simulation results of the multi-component distillation column to study the effects of feed flow rate of Bonny light crude oil on the performance of the distillation column showed that as the feed rate was increased, there was an increase in composition of lighter ends in the bottom trays (i.e inefficient separation) which was as a result of increase in velocity of the feed, decrease in residence time (contact time for liquid and vapor on each tray), and reduction in the purity of each component. On the other hand, decreasing the feed rate to a minimum led to decreased liquid and vapor flow rates hence yield was small.

These extreme flowrates result in inefficient separation, hence the need to find the optimal feed flow rate based on the most economically viable products. The graph below shows the variation of product yield for different feed rates.

Depending on the yield of light and heavy key to be achieved in the distillation column, an optimum feed flow rate is chosen.



KEY ; ***PRODUCT 1*** *– NAPHTHA,* ***PRODUCT 2*** *– KEROSENE,* ***PRODUCT 3*** *– DIESEL,*

PRODUCT 4 *- AGO,* ***PRODUCT 5*** *- RESIDUE*

Figure 4. 1 Variation of Product Yield with Feed Flowrates

4.2 Temperature Variation in the Column

The temperature distribution of the column affects the degree of separation of components of the crude oil. Each tray is to be kept at an optimum temperature to ensure maximum vapor-liquid interaction. Generally, the temperature of the column increases from top to bottom. This is because the main source of heat to the column is from the reboiler which is at the bottom of the column.

The temperature of the rectifying section of the column is controlled by the reflux flow rate while that of the stripping section is controlled by the reboiler steam flow rate.

The temperature distribution is represented in graphical format as shown below;

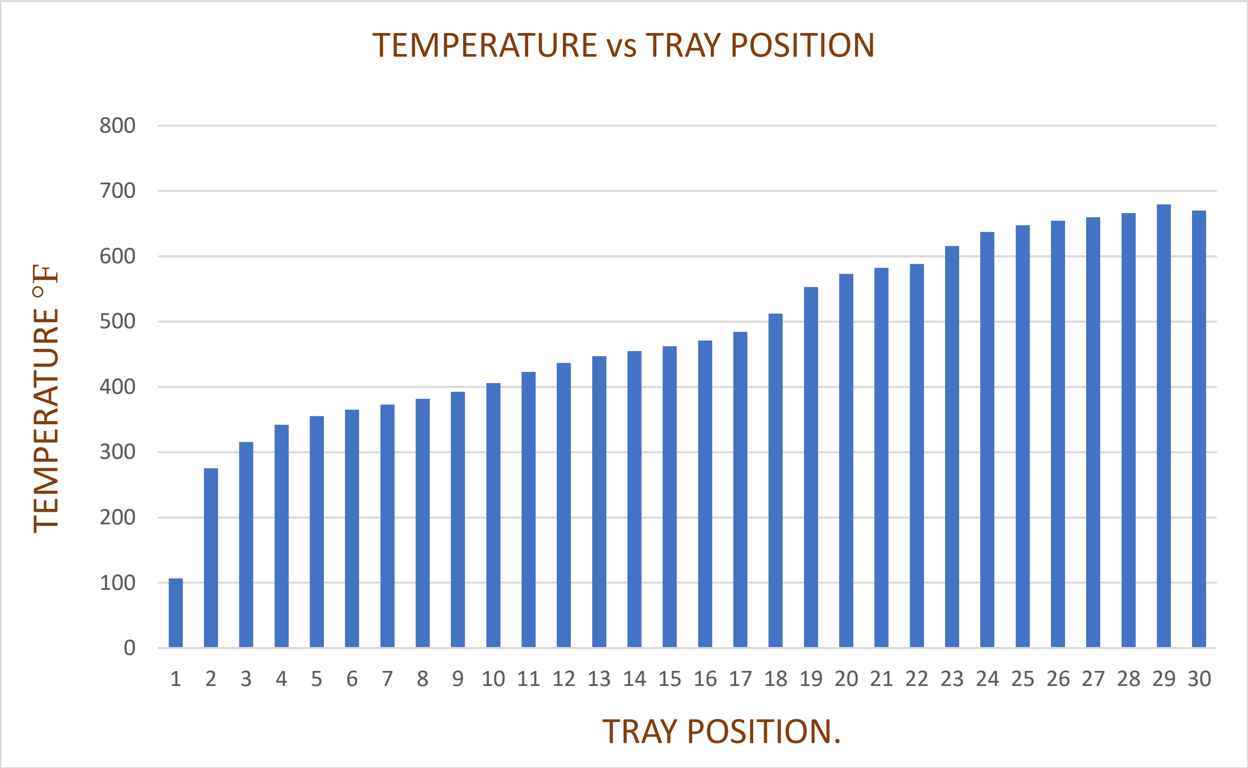


Figure 4. 2 Temperature Variation in the Column

The temperature increases steeply between tray 1 and 2 because reflux enters the column in tray 2, after which it maintains a steady increase until it gets to the stripping section. After the last tray 29, there is a drop in temperature because the bottom product has left the column.

Other control systems available include;

- Side-Products Flow

The flow of side draws into the side strippers is controlled by manipulating the flowrate of side-draw stream into the side stripper. This is to prevent overloading of the side strippers.

- Condenser Pressure

The pressure in the condenser is controlled by manipulating the flowrate of off-gas out of it. The higher the flowrate of off-gas produced, the lower the pressure in the condenser. This is done so pressure does not build up in the condenser which can lead to explosion in the plant.

- Column Feed Temperature

The temperature of the liquid feed into the column is controlled by the flowrate of fuel into the furnace so as to manipulate the air to fuel ratio for combustion. However, the feed temperature must be kept at optimum in the range of 350°C-360°C so as to reduce any disturbances that could arise as a result of temperature fluctuation.

- Series of Pre-heater Trains Temperature

The temperature of the pre-heaters 1& 2 is controlled by the flowrate of steam into the heat exchanger. This ensures the feed into the pre-flash drum is at optimum temperature.

- Level Control

The purpose of level controllers is to avoid overloading and spillage of liquid in each of the units listed below.

Condenser: Controlled by flowrate of top-product into condenser.

Side-Strippers: Controlled by flowrate of side products.

Distillation Column: Controlled by flowrate of residue stream.

4.3 Control of Level at the Diesel Chimney Tray

The choice of control of level in the diesel chimney tray is justified because diesel is the most choice product from the modular refinery. The diagrams below show the various simulations and results for the modular refinery with the presence of a PID controller. The PID controller is then tuned in Matlab- Simulink software using Ziegler Nicholas, Cohen-Coon, ITAE, and IMC methods of controller tuning to obtain the tuning method that gives the desired dynamic response to the process.

The criteria to be used in determining the tuning method with desired dynamic response is small overshoot, quick rise time, very little oscillation, and quick settling time which describes the characteristics of an underdamped system.

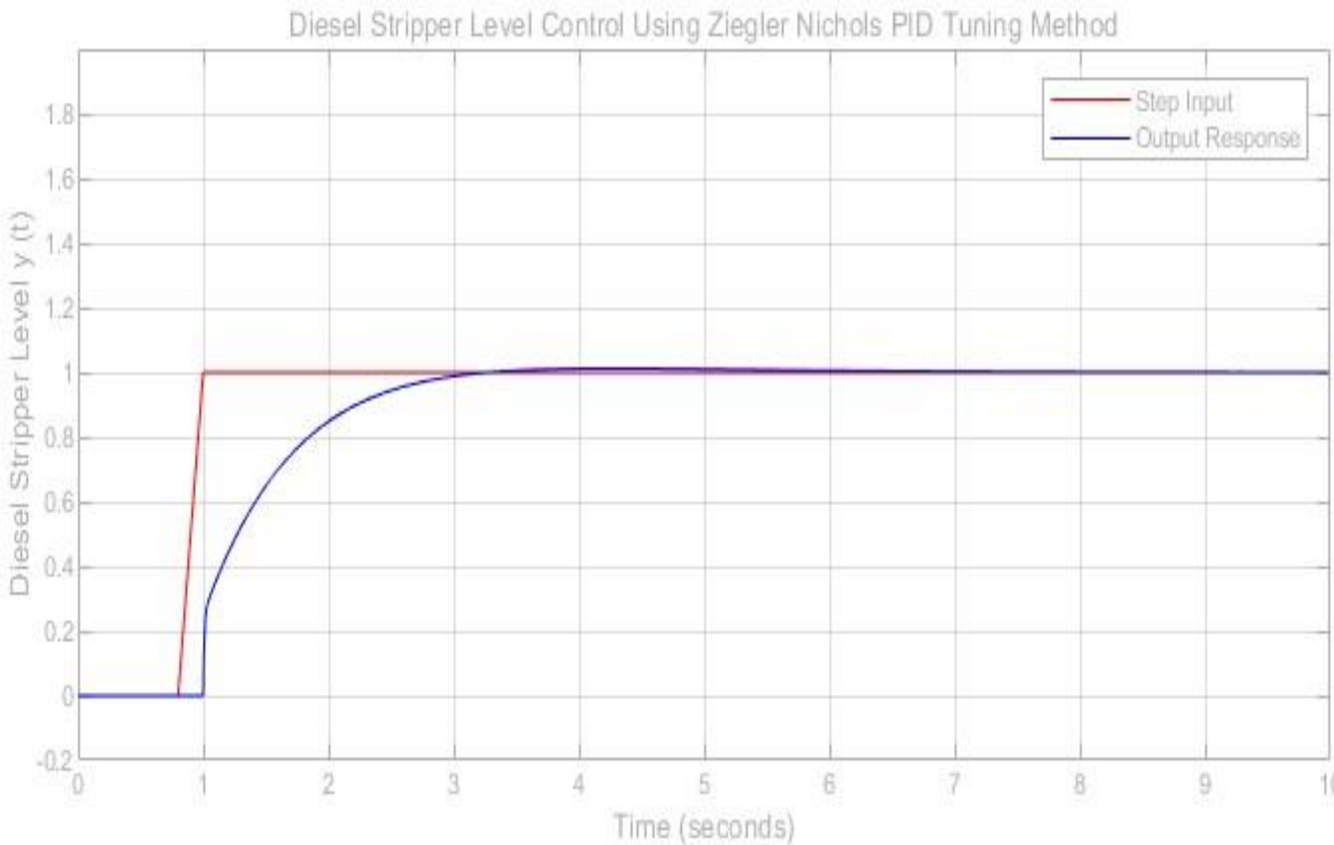


Figure 4. 3 Output Response with PID Controller Using Ziegler-Nicholas Tuning Method

The Ziegler Nicholas tuning method gave a dynamic response with approximately 1% overshoot, a rise time of 3.3s, a settling time of 7.5s, and minimal oscillation. It exhibits the characteristics of an underdamped 2^{nd} order system.

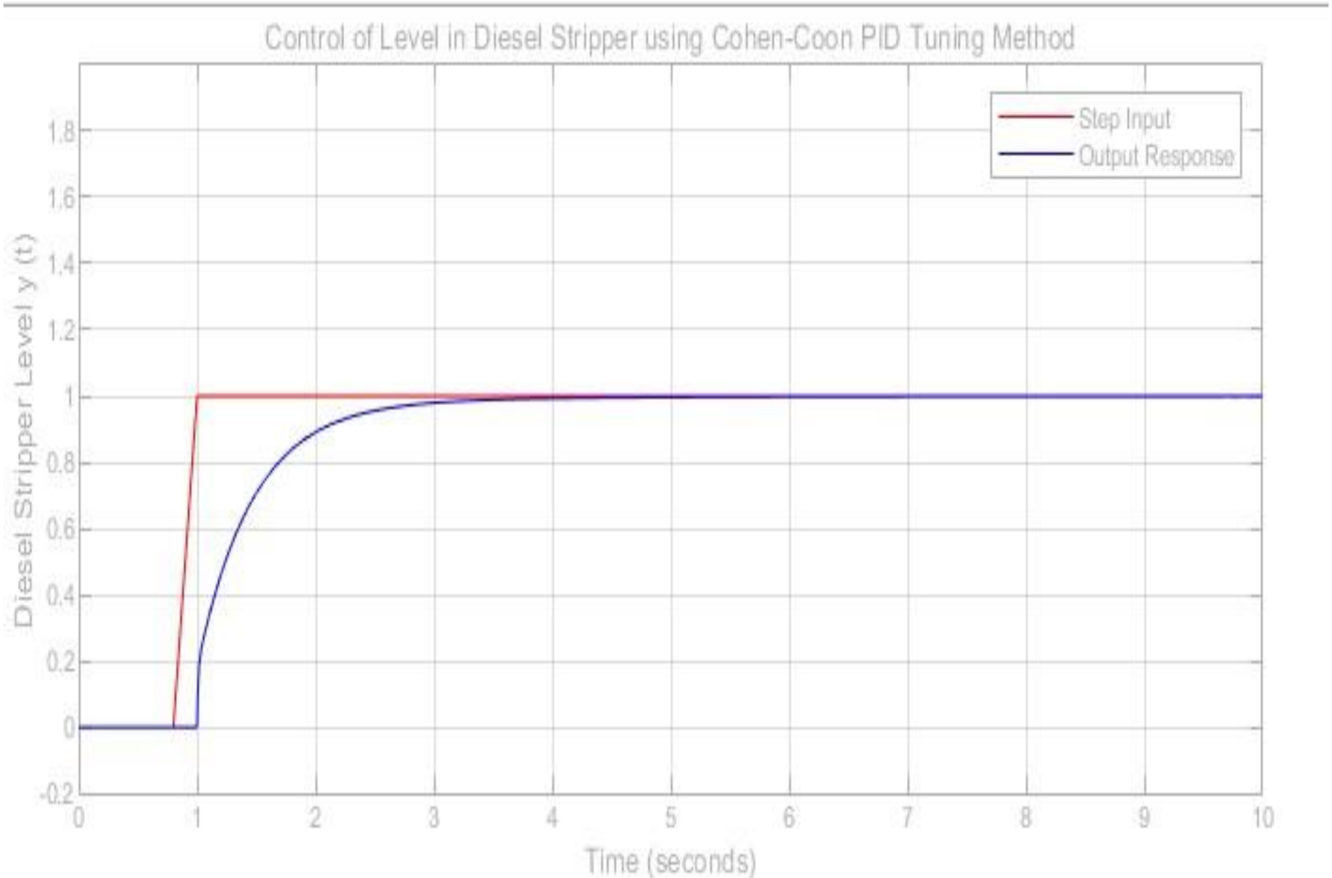


Figure 4. 4 Output Response with PID Controller Using Cohen Coon Tuning Method

The Cohen-Coon tuning method gave a dynamic response with no overshoot, a rise time of 5s, a settling time of 9s, no oscillation, and therefore behaves like a critically damped 2^{nd} order system.

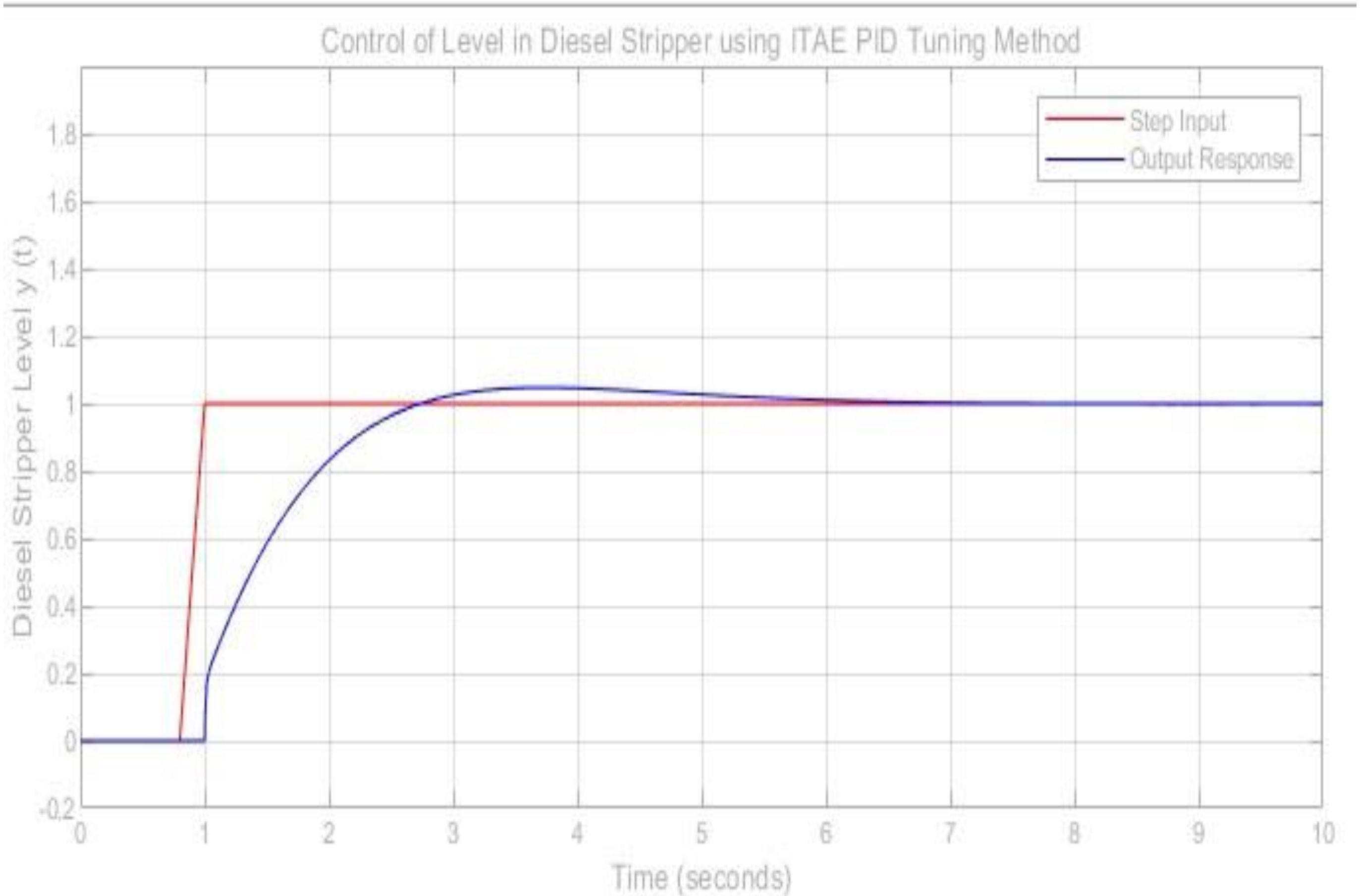


Figure 4. 5 Output Response with PID Controller Using ITAE Tuning Method

The ITAE tuning method gave a dynamic response with 4% overshoot, a rise time of 2.5s, minimal oscillation, and a settling time of 7s. It exhibits the characteristics of an underdamped 2nd order system.

The ITAE PID tuning method is the best controller tuning method for level in the diesel chimnet tray.

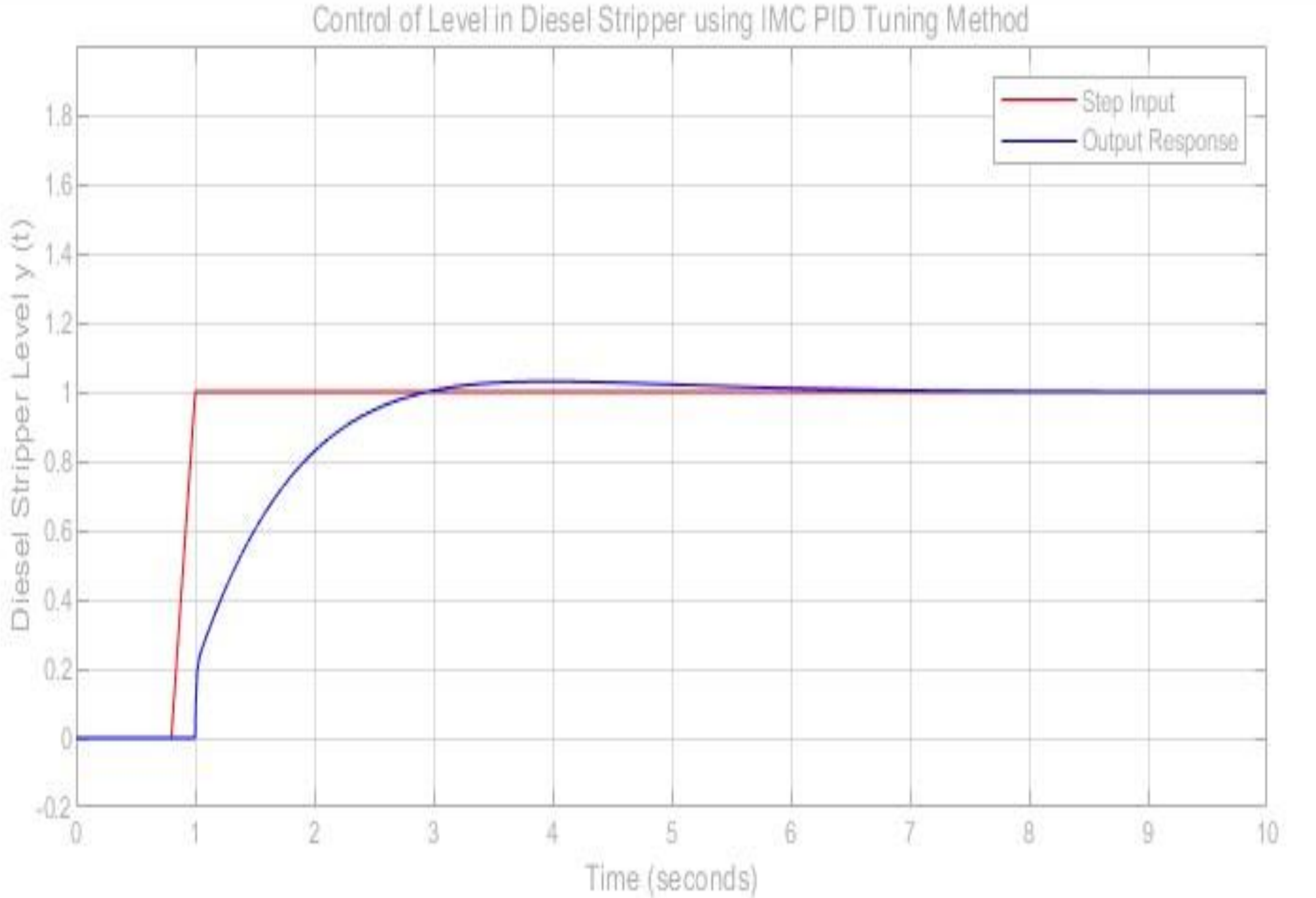


Figure 4. 6 Output Response with PID Controller Tuning Using IMC Tuning Method Method

The IMC tuning method gave a dynamic response with 3% overshoot, a rise time of 2.8s, minimal oscillation, and a settling time of 7.5s. It also exhibits the characteristics of an underdamped 2^{nd} order system.

CHAPTER FIVE – CONCLUSION AND RECOMMENDATION

CONCLUSION

In this work, a transfer function model was developed from MESH equations for a 5000 BPD modular refinery to process Bonny Light crude oil using the input-output method. The model was validated using Aspen Hysys and then simulated in MATLAB-Simulink software to control the level in the diesel chimney tray with a PID-controller tuned using the Ziegler-Nicholas, Cohen-Coon, ITAE, and IMC tuning method.

The best dynamic response was found to be the Integral Time Weighted Absolute Error (ITAE) tuning method with K_C, τ_i , and τ_D values of 2.48154, 2.14899, and 0.31699 respectively which gave a rise time of about 2.5s, 4% overshoot, minimal oscillation and a settling time of 7s thereby exhibiting the desired characteristics of an underdamped 2nd order dynamic system.

Tuning the programmable logic controller in the distributed control system of Walter-Smith modular refinery using the ITAE process parameters gotten above would minimize weeping and flooding in the column thereby ensuring maximum draw of product and maintaining temperature and pressure distribution across the tray, thus ensuring premium product quality.

RECOMMENDATION

It can be recommended that transfer function models be generated using appropriate software for future research purposes. Also, it would be helpful to also model the Naphtha tray at the top of the crude distillation column as it is the main precursor for the production of premium motor spirit or petrol, which is most consumed fuel in Nigeria.

REFERENCES

Adewuyi, D. O. (2017) 'SPE-189069-MS Flow Station Location As a Critical Factor in the Siting of a Modular Crude Oil Refinery – A Case Study of a Modular Refinery in Delta State , Nigeria', in *Society of Petroleum Engineers*, pp. 1–16. doi: SPE-189069-MS.

Ahmad, I. *et al.* (2018) 'An Artificial Intelligence Method for Energy Efficient Uncertain Feed Composition', *energies*, pp. 1–12. doi: 10.3390/en11112993.

Ahmed, D. F. and Khalaf, A. H. (2015) 'Development of Artificial Neural Network Model of Crude Oil Distillation Column', *Tikrit Journal of Engineering Sciences*, 22(1), pp. 24–37. Available at: http://www.tj-es.com.

Al-mayyahi, M. A. *et al.* (2011) 'Investigating the trade-off between operating revenue and CO 2 emissions from crude oil distillation using a blend of two crudes', *Fuel*, 90(12), pp. 3577–3585. doi: 10.1016/j.fuel.2010.12.043.

Al-mayyahi, M. A., Hoadley, A. F. A. and Rangaiah, G. P. (2014) 'Energy optimization of crude oil distillation using different designs of pre- fl ash drums', *Applied Thermal Engineering*, 73(1), pp. 1204–1210. doi: 10.1016/j.applthermaleng.2014.09.024.

Al-Salem, S. . (2015) 'ASSESSMENT OF CO2 EMISSION SOURCES FROM THE PETROLEUM SECTOR IN KUWAIT CO2 EMISSIONS IN KUWAIT FROM THE UPSTREAM', *Kuwait Institute for Scientific Research (KISR)*, pp. 1–13. doi: 10.1.1.669.3533.

Amin, M. *et al.* (2018) 'An Integrated Mechanism of Genetic Algorithm and Taguchi Method for Cut-point Temperatures Optimization of Crude Distillation Unit', in *International Conference on Computing, Mathematics and Engineering Technologies*. Islamabad, Pakistan: iCoMET 2018, pp. 1–6.

Angela, M. *et al.* (2016) 'MODULAR CRUDE OIL TOPPING REFINERY : THE TOTAL UTILIZATION OF ALL MODULAR CRUDE OIL TOPPING REFINERY : THE TOTAL UTILIZATION OF ALL DISTILLED CUTS', *Ewemen Journal of Pertrochemical Research and Innovation*, 1(2), p. a-j. Available at: http://www.researchgate.net/publication/327019521.

Angela, M. *et al.* (2019) 'Challenges and Prospects of Converting Nigeria Illegal Refineries to Modular Refineries', *The Open Chemical Engineering Journal*, 13(1), pp. 1–6. doi: 10.2174/1874123101913010001.

Bagajewicz, M. and Ji, S. (2001) *Rigorous Procedure for the Design of Conventional Atmospheric Crude Fractionation Units . Part I : Targeting*. University of Oklahoma. doi: 10.1021/ie000302.

Baldea, M. *et al.* (2017) 'Modular manufacturing processes: Status, challenges, and opportunities', *AIChE Journal*, 63(10), pp. 4262–4272. doi: 10.1002/aic.15872.

Barton, P. I. *et al.* (2015) 'Refinery Optimization Integrated with a Non-Linear Crude Distillation Unit Model', *IFAC-PapersOnLine*, 48(8), pp. 205–210. doi: 10.1016/j.ifacol.2015.08.182.

Bassem-Sayed, H. k *et al.* (2016) *Graduation Project-2 " Revamping of Crude Distillation Unit " Final Report*. Cairo.

Benali, T., Tondeur, D. and Jaubert, J. N. (2012) 'An improved crude oil atmospheric distillation process for energy integration : Part I : Energy and exergy analyses of the process when a fl ash is installed in the preheating train', *Applied Thermal Engineering*, 32, pp. 125–131. doi: 10.1016/j.applthermaleng.2011.08.038.

Chen, L. (2008) *Heat Integrated Crude Oil Distillation System design School of Chemical Engineering and Analytical Science*. The University of Manchester.

Content, L. (2018) *Modular Refineries and Local Content growth*.

Dauda, I., Megan, J. and Gonzalo, G. (2017) *Optimization-based Design of Crude Oil Distillation Units using Rigorous Simulation Models*, *American Chemical Society*. University of Manchester, Manchester M13 9PL, UK. doi: 10.1021/acs.iecr.7b01014.

Department of petroleum Resources, N. (2017) 'General Requirements and Guidance Information for the Establishment of', (May), pp. 1–5.

Elliot, S. G. . (2016) *PREHISTORIC MAN AND HISTORY*. PHILADELPHIA: J . B. LIPPIN COTT COMPAN Y. Available at: www.forgottenbooks.com.

Errico, M., Tola, G. and Mascia, M. (2009) 'Energy saving in a crude distillation unit by a preflash implementation', *Applied Thermal Engineering*, 29(8–9), pp. 1642–1647. doi: 10.1016/j.applthermaleng.2008.07.011.

Gadalla, M. *et al.* (2005) 'Reducing CO 2 Emissions and Energy Consumption of Heat-Integrated Distillation Systems', 39(17), pp. 6860–6870. doi: 10.1021/es049795q.

Gadalla, M. *et al.* (2006) 'Estimation and reduction of CO 2 emissions from crude oil distillation units', *Elsevier*, 31, pp. 2398–2408. doi: 10.1016/j.energy.2005.10.030.

Gadalla, M. A. *et al.* (2015) 'A rigorous simulation-based procedure for retrofitting an existing Egyptian refinery distillation unit', *Energy*, pp. 1–10. doi: 10.1016/j.energy.2015.02.085.

Gonçalves, D. D., Martins, F. G. and Azevedo, S. F. De (2010) 'Dynamic Simulation and Control : Application to Atmospheric Distillation Unit of Crude Oil Refinery', in Pierucci, S. and Ferraris, G. B. (eds) *20th European Symposium on Computer Aided Process Engineering – ESCAPE20*. Elsevier Ltd.

Gowlett, J. A. J. (2016) ‘The discovery of fire by humans : a long and convoluted process’, *Philosophical Transactions of The Royal Society B Biological Sciences*, (May), pp. 1–12. doi: 10.1098/rstb.2015.0164.

Handogo, R. (2012) ‘Optimization on the performance of crude distillation unit (CDU)’, *ASIA-PACIFIC JOURNAL OF CHEMICAL ENGINEERING*, 7(November 2011), pp. 78–87. doi: 10.1002/apj.

Hogan, J. S. (1976) ‘United States Patent (19)’. United States of America. Available at: www.scholar.google.com (Accessed: 27 February 2020).

Idris, M. N. *et al.* (2018) ‘Design and Development of 15,000 Barrel per day (BPD) Capacity of Modular Crude Oil Refinery Plant’, *International Journal of Engineering and Modern Technology*, 4(2), pp. 1–13.

Ji, S. and Bagajewicz, M. (2002) ‘Design of Crude Fractionation Units with Preflashing or Prefractionation : Energy Targeting’, *American Chemical Society*, pp. 3003–3011. doi: 10.1021/ie010509i.

Kandasamy, V. *et al.* (2019) ‘Optimizing atmospheric distillation unit for maximum light petroleum gas yield and comparative case studies Optimizing Atmospheric Distillation Unit for Maximum Light Petroleum Gas Yield and Comparative Case Studies’, 020050(July), pp. 1–14. doi: 10.1063/1.5117110.

Ledezma-martínez, M., Jobson, M. and Smith, R. (2018) *Simulation-optimization-based Design of Crude Oil Distillation Systems with Preflash Units*. The University of Manchester, Manchester M13 9PL, UK. doi: 10.1021/acs.iecr.7b05252.

Liebmann, K., Dhole, V. R. and Jobson, M. (1998) ‘DISTILLATION TOWER USING PINCH ANALYSIS’, *Institution of Chemical Engineers, IChemE*, 76(March), pp. 335–337.

Lopez, D.-C. *et al.* (2009) ‘OPTIMIZATION MODEL’, 3(5), pp. 159–174.

Mamudu, A. O., Igwe, G. J. and Okonkwo, E. (2019) ‘Process design evaluation of an optimum modular topping refinery for Nigeria crude oil using hysys ® aspen software Process design evaluation of an optimum modular topping refinery for Nigeria crude oil using hysys ® aspen software’, *Cogent Engineering*, 1916, pp. 1–17. doi: 10.1080/23311916.2019.1659123.

Mamudu, O. A. *et al.* (2016) ‘Process Design Evaluation of an Optimum Modular Topping Refinery for Nigeria Crude Oil Using HYSYS Software’, in *Society of Petroleum Engineers*. Port-Harcourt: Society of Petroleum Engineers, pp. 1–16.

Minh, V. T. and Pumwa, J. (2012) ‘Modeling and Control Simulation for a Condensate Distillation Column’, *Research Gate*, (March), pp. 1–33. doi: 10.5772/38651.

Morgan, K. (2017) 'The need for fossil fuel', in *Fossil Fuel*, pp. 1–19. Available at: www.kindermorgan.com.

National Programme on Technology Enhanced Learning (2006) 'Lecture 3 : Petroleum Refining Overview', in *Petroleum Refining*, pp. 1–83.

Ogbon, N. O., Otanocha, O. B. and Rim-Rukeh, A. (2018) 'An assessment of the economic viability and competitiveness of modular refinery in Nigeria', *Nigerian Journal of Technology*, 37(4), p. 1015. doi: 10.4314/njt.v37i4.22.

Rastogi, V. (2006) *Heat-integrated crude oil distillation system design*. The University of Manchester.

Serdar, K. *et al.* (2004) 'MODEL PREDICTIVE CONTROL OF A CRUDE DISTILLATION UNIT'.

Sharma, R. *et al.* (1999) 'Design / Retrofit Targets of Pump-Around Refluxes for Better Energy Integration of a Crude Distillation Column', *American Chemical Society*, pp. 2411–2417. doi: 10.1021/ie9802156.

Shigueo, A., Carlos, A. and Odloak, D. (2015) 'Tuning the Model Predictive Control of a Crude Distillation Unit', *ISA Transactions*, pp. 1–13. doi: 10.1016/j.isatra.2015.10.017.

Stephanopoulous, G. (1983) *Chemical Process Control An Introduction to Theory and Practice Problem Sotving in Chemical Engineering with Numerical Methods Fluid Mechanics for Chemical Engineering*. Edited by A. NEALR. Athens, Greece: PRENTICE HALL, Englewood Cliffs, New Jersey.

Straelen, J. Van *et al.* (2009) 'Energy Procedia CO 2 capture for refineries , a practical approach', *Energy Procedia*, 1(1), pp. 179–185. doi: 10.1016/j.egypro.2009.01.026.

Suphanit, B. and Smith, R. (1999) *Variable print quality*. University of Manchester Institute of Science and Technology.

Syed, S. P. (2014) 'Fuels and Combustion', pp. 1–82.

UNITED STATES DEPARTMENT OF COMMERCE, B. O. S. (1929) *THERMAL PROPERTIES OF PETROLEUM PRODUCTS*. UNITED STATE GOVERNMENT PRINTING OFFICE, WASHINGTON.

Wanders, T. (2016) 'Reducing CO2 emissions of the Dutch refining industry towards 2050', 2009, pp. 1–14.

Wang, Y. *et al.* (2011) 'Selecting the Optimum Predistillation Scheme for Heavy Crude Oils', *Industrial and Engineering Chemistry Research*, pp. 10549–10556. doi: 10.1021/ie102369k.

Watkins, R. N. (1979) *Watkins, R. - 1979 - Petroleum Refinery Distillation.pdf*. Houston Texas: Gulf Publishing Company.

Wrigley, E. A. (2015) ‘Energy and the English industrial revolution’, *Energy and the English Industrial Revolution*, pp. 1–272. doi: 10.5860/choice.48-4603.

Yavini, T. D., Namo, J. M. and Sani, I. L. (2015) ‘Development of Optimal Strategy for the Design and Operation of a Crude Petroleum Distilation (Topping) Unit Price of Gas Oil) − ∑ n’, *Journal of Bioprocessing & Chemical Engineering*, 3(1), pp. 1–4. doi: 10.15297/JBCE.V3I1.01.

Zhang, J., Zhu, X. X. and Towler, G. P. (2001) ‘A Simultaneous Optimization Strategy for Overall Integration in Refinery Planning’, *American Chemical Society*, pp. 2640–2653. doi: 10.1021/ie000367c.

Zhang, N. and Zhu, X. X. (2000) ‘Computers & Chemical Engineering A novel modelling and decomposition strategy for overall refinery optimisation’, *American Chemical Society*, 24, pp. 1543–1548.