

**DESIGN OF AN ATMOSPHERIC DISTILLATION UNIT
CONFIGURATION TO REFINE 4800 m³/day OF CRUDE OIL**

By

GROUP 15

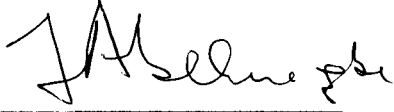
**A DESIGN PROJECT SUBMITTED TO THE
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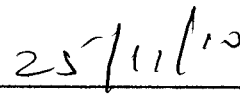
CERTIFICATION

This is to certify that this project report (thesis) entitled 'DESIGN OF AN ATMOSPHERIC DISTILLATION UNIT CONFIGURATION TO REFINE 4800 m³/day OF CRUDE OIL' by Group 15 meets the requirements for the partial fulfilment of the award of Bachelor of Engineering (B.Eng) degree in Chemical Engineering, Federal University of Technology, Minna.

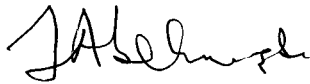


Project supervisor

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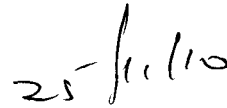


Date



Head of Department

Name and Signature



Date



External Examiner

Name and Signature

Prof. Dr. P. B. OJIJI



Date

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ABSTRACT

This design project was aimed at designing an atmospheric distillation unit configuration to refine 4800 m³/day of crude oil. The atmospheric distillation column was designed with a Preflash implementation technique which enhanced efficient energy utilization of the whole refining process. A total capital investment of 3,540,843,138 Naira was required with a payback period and rate of return of 5 years and 20 % respectively.

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GENERAL INTRODUCTION

The entire work of this design project entitled Design of an Atmospheric Distillation Unit Configuration to Refine 4800 m³/day of Crude Oil was jointly carried out by ALL members of group 15. Below is a summary of how the work was carried out by members of the group.

Chapters One and Two were compiled by all members of the group.

The Material and Energy balance (i.e. Chapters Three and Four) were carried out by Arungwa James (2005/21726EH), Arijesuyo Gabriel O. (2006/25839EH) and Abdulrasheed Sherif (2006/24149EH).

Chapters Six and Seven (Equipment Design and Optimization) were done by Arungwa James (2005/21726EH).

Chapter Eight (Safety and Quality Control) was done by Mathew Ikenna Francis (2006/24140EH) and Salihu Bello (2006/25772EH).

Chapter Nine (Process Control and Instrumentation) was done by Adegoke Seun (2005/21563EH).

Chapter Ten (Environmental Acceptability) was jointly carried out by Olumi Olufunke (2005/21783EH) and Peter Idongesit Akpan (2005/21793EH).

Chapter Eleveen (Start-up and Shutdown) was carried out by Arungwa James (2005/21726EH).

Chapter Twelve (Site for Plant Location) was mutually done by Buzu Ezekiel Garba (2005/21730EH) and Odeh Onoja Junoir (2005/21774EH).

Chapter Thirteen (Economic Analysis) was done by Abdulrasheed Sherif (2006/24149EH) and Arijesuyo Gabriel O. (2006/25839EH).

Chapter Fourteen (Conclusions and Recommendations) was done by all members of the group.

GROUP MEMBERS.

Matric. Number	Name	Designation
2005/21726EH	Arungwa, James Chikwendu	Leader
2006/25839EH	Arijesuyo, Gabriel O.	Asst. Leader
2006/24149EH	Sherif, Abdulrasheed	Asst. Leader
2005/21730EH	Buzu, Ezekiel Garba	Member
2006/24140EH	Mathew, Ikenna Francis	Member
2005/21774EH	Odeh, Onoja Junoir	Member
2005/21783EH	OLUMI, OLUFUNKE	Member
2005/21793EH	PETER, IDONGESIT AKPAN	Member
2006/25772EH	SALIHU, BELLO SALIHU	Member
2005/21563EH	SEUN, ADEGOKE	Member

Chapter One

1.0 INTRODUCTION

Petroleum is a fossil fuel. Petroleum is often called crude oil, or oil. It is called fossil fuel because it was formed from the remains of tiny sea plants and animals that died millions of years ago. When the plants and animals died, they sank to the bottom of the oceans. Here, they were buried by thousands of feet of sand and sediment, which turned into sedimentary rock. As the layers increased, they pressed harder and harder on the decayed remains at the bottom. The heat and pressure changed the remains, and eventually, petroleum was formed. Petroleum deposits are locked in porous rocks almost like water is trapped in a wet sponge. When crude oil comes out of the ground, it can be as thin as water or as thick as tar. Petroleum is called nonrenewable energy source because it takes millions of years to form. We cannot make new petroleum reserves (Wikipedia, 2010).

In fact, modern industrial civilization depends on petroleum and its products; the physical structure and way of life of the suburban communities that surround the great cities are the result of an ample and inexpensive supply of petroleum. In addition, the goals of developing countries—to exploit their natural resources and to supply foodstuffs for the burgeoning populations—are based on the assumption of petroleum availability. In recent years, however, the worldwide availability of petroleum has steadily declined and its relative cost has increased. Many experts forecast that petroleum will no longer be a common commercial material by the mid-21st century (Wikipedia, 2010).

People have used petroleum since ancient times. The ancient Chinese and Egyptians burned oil to light their homes. Before the 1850s, Americans used whale oil to light their homes. When whale oil became scarce, people skimmed the oil that seeped to the surface of ponds and streams. The demand for oil grew, and in 1859, Edwin Drake drilled the first oil well near Titusville, Pennsylvania. At first, the crude oil was refined or made into kerosene for lighting. Gasoline and other products made during refining were thrown away because people had no use for them. This all changed when Henry Ford began mass production of automobiles in 1913. Everyone wanted an automobile and they all ran on gasoline. Gasoline was the fuel of choice because it provided the greatest amount of energy in

relation to cost and ease of use. Today, Americans use more petroleum than any other energy source, mostly for transportation. Petroleum provides almost 39 percent of the energy we use.

An oil refinery or petroleum refinery is an industrial process plant where crude oil is processed and refined into more useful petroleum products, such as gasoline, diesel fuel, asphalt base, heating oil, kerosene, and liquefied petroleum gas (Gary, 1984). Oil refineries are typically large sprawling industrial complexes with extensive piping running throughout, carrying streams of fluids between large chemical processing units. In many ways, oil refineries use much of the technology of, and can be thought of as types of chemical plants. The crude oil feedstock has typically been processed by an oil production plant. There is usually an oil depot (tank farm) at or near an oil refinery for storage of bulk liquid products (Alphonsus, 1991).

Oil refineries are large scale plants, processing about a hundred thousand to several hundred thousand barrels of crude oil a day. Because of the high capacity, many of the units operate continuously, as opposed to processing in batches, at steady state or nearly steady state for months to years. The high capacity also makes process optimization and advanced process control very desirable (Heather, 2009).

Once separated and purified of any contaminants and impurities, the fuel or lubricant can be sold without further processing. Smaller molecules such as isobutane and propylene or butylenes can be recombined to meet specific octane requirements by processes such as alkylation, or less commonly, dimerization. Octane grade of gasoline can also be improved by catalytic reforming, which involves removing hydrogen from hydrocarbons producing compounds with higher octane ratings such as aromatics. Intermediate products such as gas oils can even be reprocessed to break a heavy, long-chained oil into a lighter short-chained one, by various forms of cracking such as fluid catalytic cracking, thermal cracking, and hydrocracking. The final step in gasoline production is the blending of fuels with different octane ratings, vapor pressures, and other properties to meet product specifications (Heather, 2009).

1.1 Aim and Objective

The aim of this work is to design an Atmospheric distillation configuration to refine crude oil into straight-run products. The aim is to be achieved by designing the Configuration with Preflash column implementation.

1.2 Scope of Work

This design project is limited only to the refining of a pre-treated crude oil into straight run products using an atmospheric distillation unit.

1.3 Justification

Considering Nigeria's position as one of the world leading oil producing Nations as well as having one of the finest oil deposits, with inadequate refineries, and also given the fact that the oil sector is a very crucial sector in the Nigeria economy, there is the dire need for more refineries to produce appropriate and desirable products basically for domestic consumption and to also increase the total GDP of the economy.

Chapter Two

2.0 LITERATURE REVIEW

Petroleum, or crude oil, naturally occurring oily, bituminous liquid composed of various organic chemicals. It is found in large quantities below the surface of Earth and is used as a fuel and as a raw material in the chemical industry. Modern industrial societies use it primarily to achieve a degree of mobility—on land, at sea, and in the air—that was barely imaginable less than 100 years ago. In addition, petroleum and its derivatives are used in the manufacture of medicines and fertilizers, foodstuffs, plastics, building materials, paints, and cloth and to generate electricity (Crude Oil-wikipedia.com, 2010).

The first oil refineries in the world were built by Ignacy Łukasiewicz near Jasło, Austrian Empire (now in Poland) from 1854 to 1856 (Frank and Alison, 2005) but they were initially small as there was no real demand for refined fuel. As Łukasiewicz's kerosene lamp gained popularity, the refining industry grew in the area.

World's first large refinery opened at Ploesti (today known as Ploiești), Romania, in 1856-1857 (World Events: 1844-1856), with US investment. After being taken over by Nazi Germany, the Ploesti refineries were bombed in Operation Tidal Wave by the Allies during the Oil Campaign of World War II. Another early large refinery is Oljeön, Sweden (1875), now preserved as a museum at the UNESCO world heritage site Engelsberg and part of the Ecomuseum Bergslagen (Wikipedia, 2010).

At one point, the refinery in Ras Tanura, Saudi Arabia owned by Saudi Aramco was claimed to be the largest oil refinery in the world. For most of the 20th century, the largest refinery was the Abadan Refinery in Iran. This refinery suffered extensive damage during the Iran-Iraq war. The world's largest refinery complex is the Jamnagar Refinery Complex, consisting of two refineries side by side operated by Reliance Industries Limited in Jamnagar, India with a combined production capacity of 1,240,000 barrels per day (197,000 m³/d) (J-1 660,000 bbl/d (105,000 m³/d), J-2 580,000 bbl/d (92,000 m³/d).

PDVSA's Paraguana refinery complex in Venezuela with a capacity of 956,000 bbl/d (152,000 m³/d) and SK Energy's Ulsan in South Korea with 840,000 bbl/d (134,000 m³/d) are the second and third largest, respectively (Erik, 2003).

These surface deposits of crude oil have been known to humans for thousands of years. In the areas where they occurred, they were long used for limited purposes, such as caulking boats, waterproofing cloth, and fueling torches. By the time the Renaissance began in the 14th century, some surface deposits were being distilled to obtain lubricants and medicinal products, but the real exploitation of crude oil did not begin until the 19th century. The Industrial Revolution had by then brought about a search for new fuels, and the social changes it effected had produced a need for good, cheap oil for lamps; people wished to be able to work and read after dark. Whale oil, however, was available only to the rich, tallow candles had an unpleasant odor, and gas jets were available only in then-modern houses and apartments in metropolitan areas (Erik, 2003).

The search for a better lamp fuel led to a great demand for “rock oil”—that is, crude oil—and various scientists in the mid-19th century were developing processes to make commercial use of it. Thus British entrepreneur James Young, with others, began to manufacture various products from crude oil, but he later turned to coal distillation and the exploitation of oil shales. In 1852 Canadian physician and geologist Abraham Gessner obtained a patent for producing from crude oil a relatively clean-burning, affordable lamp fuel called kerosene; and in 1855 an American chemist, Benjamin Silliman, published a report indicating the wide range of useful products that could be derived through the distillation of petroleum (Erik, 2003).

Thus the quest for greater supplies of crude oil began. For several years people had known that wells drilled for water and salt were occasionally infiltrated by petroleum, so the concept of drilling for crude oil itself soon followed. The first such wells were dug in Germany from 1857 to 1859, but the event that gained world fame was the drilling of an oil well near Oil Creek, Pennsylvania, by “Colonel” Edwin L. Drake in 1859. Drake,

contracted by the American industrialist George H. Bissell—who had also supplied Silliman with rock-oil samples for producing his report—drilled to find the supposed “mother pool” from which the oil seeps of western Pennsylvania were assumed to be emanating. The reservoir Drake tapped was shallow—only 21.2 m (69.5 ft) deep—and the petroleum was a paraffin type that flowed readily and was easy to distill.

Drake’s success marked the beginning of the rapid growth of the modern petroleum industry. Soon petroleum received the attention of the scientific community, and coherent hypotheses were developed for its formation, migration upward through the earth, and entrapment. With the invention of the automobile and the energy needs brought on by World War I (1914-1918), the petroleum industry became one of the foundations of industrial society (Trambouze, 2001).

In the 1800s, refineries in the U.S. processed crude oil primarily to recover the kerosene. There was no market for the more volatile fraction, including gasoline, which was considered waste and was often dumped directly into the nearest river. The invention of the automobile shifted the demand to gasoline and diesel, which remain the primary refined products today. Today, national and state legislation requires refineries to meet stringent air and water cleanliness standards. In fact, oil companies in the U.S perceive obtaining a permit to build a modern refinery to be so difficult and costly that no new refineries have been built (though many have been expanded) in the U.S since 1976. More than half the refineries that existed in 1981 are now closed due to low utilization rates and accelerating mergers. As a result of these closures, total US refinery capacity fell between 1981 to 1995. In 2009 through 2010, as revenue streams in the oil business dried up and profitability of oil refineries fell due to lower demand for product and high reserves of supply preceding the economic recession, oil companies began to close or sell refineries. Due to EPA regulations, the costs associated with closing a refinery are very high, meaning that many former refineries are repurposed (Trambouze, 2001).

2.1 Types of Crude Oil

Crude oils are complex mixtures containing many different hydrocarbon compounds that vary in appearance and composition from one oil field to another. Crude oils range in consistency from water to tar-like solids, and in color from clear to black. An average crude oil contains about 84 percent carbon, 14 percent hydrogen, 1 to 3 percent sulphur, and less than 1 percent each of nitrogen, oxygen, metals, and salts. Crude oils are generally classified as paraffinic, naphthenic, or aromatic based on the predominant proportion of similar hydrocarbon molecules. Mixed-base crudes have varying amounts of each type of hydrocarbon. Refinery crude base stocks usually consist of mixtures of two or more different crude oils. Table 2.1 lists some typical properties for crude oil sources from around the world (Wikipedia, 2010).

Crude oils are also defined in terms of API (American Petroleum Institute) gravity number. The higher the API gravity number, the lighter the crude. For example, light crude oils have high API gravities and low specific gravities. Crude oils with low carbon, high hydrogen, and high API gravity are usually rich in paraffins and tend to yield greater proportions of gasoline and light petroleum products. Crude oils with high carbon, low hydrogen, and low API gravities are usually rich in aromatics. Crude oils that contain appreciable quantities of hydrogen sulfide or other reactive sulphur compounds are called sour. Those with less sulphur are called sweet. Some exceptions to this rule are the West Texas crudes, which are always considered sour regardless of their H₂S content, and the Arabian high-sulphur crudes, which are not considered sour because their sulphur compounds are not highly reactive.

For refining operations, the acidity of the crude oil is an important consideration for economic reasons. A number of organic acids may be present in crude oil feedstocks. The extra costs associated with handling high-acid crudes can be offset by a lower feedstock cost. Acidity is defined in terms of the total acid number (TAN), which is a measure of the number of milligrams of potassium hydroxide (KOH) needed to neutralize 1 g of sample. A TAN exceeding 1.5 to 1.8 mg KOH/g is considered corrosive; however, corrosion problems can occur in crudes with

TAN numbers as low as 0.3 for several reasons, including velocity and the nature of the acidic species present.

Table 2.1: Typical approximate characteristics, properties and gasoline potential of various crudes (Howard, 1987).

CRUDE SOURCE	PARAFINS (% VOL)	AROMATICS (% VOL)	NAPHTHENES (% VOL)	SULPHUR (% WT)	API GRAVITY (APPROX.)	NAPH. YIELD (% VOL)	OCTANE NUMBER (TYPICAL)
Nigerian (Light)	37	9	54	0.2	36	28	60
Saudi (Light)	63	19	18	2.0	34	22	40
Saudi (Heavy)	60	15	25	2.1	28	23	35
Venezuela (Heavy)	35	12	53	2.3	30	2	60
Venezuela (Light)	52	14	34	1.5	24	18	50
USA Midcont. Sweet	-	-	-	0.4	40	-	-
USA (W. Texas Sour)	46	22	32	1.9	32	33	55
North sea (Brent)	50	16	34	0.4	37	31	50

2.2 Composition and Characteristics of Crude Petroleum

2.2.1 General Review

Crude oils are complex mixtures of a vast number of hydrocarbon compounds. Properties of crude petroleum vary appreciably and depend mainly on the origin. In this chapter, the chemical composition of the crude oils is viewed, including the hydrocarbon series as well as the non-hydrocarbon compounds. Physical methods generally used for indentifying types of crude oils are described next. Characterization and classification of crude oils based on correlation indexes and crude assays are presented, followed by a comparison between some of the well-known types of oil.

2.2.2 Crude Oil Composition

The raw material that we deal with—referred to as crude petroleum—is, by definition, the naturally occurring rock oil produced as was explained in Chapter 1. In general, composition of crude oil may be studied by two methods:

1. Chemical approach
2. Physical methods

Chemical composition describes and identifies the individual chemical compounds isolated from crude oils over the years. Physical representation, on the other hand, involves considering the crude oil and its products as mixtures of hydrocarbons and describing physical laboratory tests or methods for characterizing their quality (Wikipedia, 2010).

Table 2.2: Composition of Crude Oil (Wikipedia, 2010)

ELEMENT	PERCENT BY WEIGHT
Carbon	83-87
Hydrogen	11-14
Sulphur	0.05-2.5
Nitrogen	0.1-2
Oxygen	0-2

2.2.2.1 Chemical Approach

Nearly all petroleum deposits are made up of a mixture of chemical compounds that consist of hydrogen and carbon, known as hydrocarbons, with varying amounts of non-hydrocarbons containing S, N₂, O₂, and other some metals. The composition of crude oil by elements is approximated as shown in Table 1. It could be further stated that these hydrocarbon compounds making up oils are grouped chemically into different series of compounds described by the following characteristics. Each series consists of compounds similar in their molecular structure and properties (e.g., the alkanes or paraffin series). Within a given series, there exists a wide spectrum of compounds that range from

extremely light or simple hydrocarbon to a heavy or complex one. An example, CH_4 for the former and $\text{C}_{40}\text{H}_{82}$ for the latter in the paraffinic series (Wikipedia, 2010).

Hydrocarbon Series:

The major constituents of most crude oils and its products are hydrocarbon compounds, which are made up of hydrogen and carbon only. These compounds belong to one of the following subclasses:

1. **Alkanes or Paraffins:** Alkanes are saturated compounds having the general formula $\text{C}_n\text{H}_{2n+2}$. Alkanes are relatively nonreactive compounds in comparison to other series. They may either be straight-chain or branched compounds, the latter are more valuable than the former, because they are useful for the production of high-octane gasoline.

2. **Cycloalkanes or Cycloparaffins (Naphthenes):** Cycloalkanes and bicycloalkanes are normally present in crude oils and its rations in variable proportions. The presence of large amounts of these cyclic compounds in the naphtha range has its significance in the production of aromatic compounds. Naphtha cuts with a high percentage of naphthenes would make an excellent feedstock for aromatization.

3. **Alkenes or Olefins:** Alkenes are unsaturated hydrocarbon compounds having the general formula C_nH_n . They are practically not present in crude oils, but they are produced during processing of crude oils at high temperatures.

Alkenes are very reactive compounds. Light olefinic hydrocarbons are considered the base stock for many petrochemicals. Ethylene, the simplest alkene, is an important monomer in this regard. For example, polyethylene is a well-known thermoplastic polymer and polybutadiene is the most widely used synthetic rubber.

4. **Aromatics:** Aromatic compounds are normally present in crude oils. Only monomolecular compounds in the range of $\text{C}_6\text{--C}_8$ (known as B-T-X) have gained commercial importance.

Aromatics in this range are not only important petrochemical feedstocks but are also valuable for motor fuels.

5. **Dinuclear and polynuclear aromatic compounds** are present in heavier petroleum fractions and residues. Asphaltene, which are concentrated in heavy residues and in some

asphaltic crude oils, are, in fact, polynuclear aromatics of complex structures. It has been confirmed by mass spectroscopic techniques that condensed-ring aromatic hydrocarbons and heterocyclic compounds are the major compounds of asphaltenes.

Non-hydrocarbon Compounds

Other types of non-hydrocarbon compound occur in crude oils and refinery streams. Most important are the following:

- Sulphur compounds
- Nitrogen compounds
- Oxygen compounds
- Metallic compounds

Sulphur Compounds:

In addition to the gaseous sulphur compounds in crude oil, many sulphur compounds have been found in the liquid phase in the form of organosulphur. These compounds are generally not acidic.

Sour crude oils are those containing a high percentage of hydrogen sulfide. However, many of the organic sulphur compounds are not thermally stable, thus producing hydrogen sulfide during crude processing.

High-sulphur crude oils are in less demand by refineries because of the extra cost incurred for treating refinery products. Naphtha feed to catalytic reformers is hydro-treated to reduce sulphur compounds to very low levels (1 ppm) to avoid catalyst poisoning.

The following sulphur compounds are typical:

1. **Mercaptans (H-S-R):** Hydrogen sulfide, H-S-H, may be considered as the simple form of mercaptan; however, the higher forms of the series are even more objectionable in smell. For example, butyl mercaptan (H-S-C₄H₉) is responsible for the unusual odor of the shank.
2. **Sulfides (R-S-R):** When an alkyl group replaces the hydrogen in the sulphur-containing molecule, the odor is generally less obnoxious. Sulfides could be removed by the hydrotreating technique, which involves the hydrogenation of the

petroleum streams as follows: The hydrogen sulfide may be removed by heating and may be separated by using amine solutions.

3. **Polysulfides (R-S-S-R):** These are more complicated sulphur compounds and they may decompose, in some cases depositing elemental sulphur. They may be removed from petroleum fractions, similar to the sulfides, by hydro-treating (Indiamart, 2010).

Nitrogen Compounds:

Nitrogen compounds in crude oils are usually low in content (about 0.1–0.9 %) and are usually more stable than sulphur compounds. Nitrogen in petroleum is in the form of heterocyclic compounds and may be classified as basic and nonbasic. Basic nitrogen compounds are mainly composed of pyridine homologs and have the tendency to exist in the high-boiling fractions and residues. The nonbasic nitrogen compounds, which are usually of the pyrrole and indole, also occur in high-boiling fractions and residues. Only a trace amount of nitrogen is found in light streams. During hydrotreatment (hydrodesulphurization) of petroleum streams, hydrodenitrogenation takes place as well, removing nitrogen as ammonia gas, thus reducing the nitrogen content to the acceptable limits for feedstocks to catalytic processes. It has to be stated that the presence of nitrogen in petroleum is of much greater significance in refinery operations than might be expected from the very small amounts present. It is established that nitrogen compounds are responsible for the following:

1. Catalyst poisoning in catalytic processes
2. Gum formation in some products such as domestic fuel oils.

Oxygen Compounds:

Oxygen compounds in crude oils are more complex than sulphur compounds. However, oxygen compounds are not poisonous to processing catalysts. Most oxygen compounds are weakly acidic, such as phenol, cresylic acid and naphthenic acids. The oxygen content of petroleum is usually less than 2 %, although larger amounts have been reported (Indiamart, 2010).

Metallic Compounds:

Many metals are found in crude oils; some of the more abundant are sodium, calcium, magnesium, iron, copper, vanadium, and nickel. These normally occur in the form of inorganic salts soluble in water—as in the case of sodium chloride—or in the form of organometallic compounds—as in the case of iron, vanadium, and nickel.

The occurrence of metallic constituents in crude oils is of considerably greater interest to the petroleum industry than might be expected from the very small amounts present. The organometallic compounds are usually concentrated in the heavier fractions and in crude oil residues. The presence of high concentration of vanadium compounds in naphtha streams for catalytic reforming feeds will cause permanent poisons. These feeds should be hydrotreated not only to reduce the metallic poisons but also to desulphurize and denitrogenate the sulphur and nitrogen compounds.

Hydrotreatment may also be used to reduce the metal content in heavy feeds to catalytic cracking (Indiamart, 2010).

2.2.2.2 Physical Methods

Due to the various chemicals found in crude oils and the complexity of the mixture with the difficulty of specifying a crude oil as a particular mixture of chemicals, early producers of petroleum adopted the physical methods generally used for classification.

As may be seen, crude oils from different locations may vary in appearance and viscosity and also vary in their usefulness as producers for final products. It is possible by the use of certain basic tests to identify the quality of crude oil stocks. The tests included in the following list are primarily physical (except sulphur determination):

1. Distillation
2. Density, specific gravity, and API (American Petroleum Industry) gravity
3. Viscosity
4. Vapor pressure
5. Flash and fire points
6. Cloud and pour points
7. Color

8. Sulphur content
9. Basic sediments and water (B.S.&W.)
10. Aniline point
11. Carbon residue

API Gravity:

Earlier, density was the principal specification for petroleum products. However, the derived relationships between the density and its fractional composition were only valid if they were applied to a certain type of petroleum. Density is defined as the mass of a unit volume of material at a specified temperature. It has the dimensions of grams per cubic centimeter. Another general property, which is more widely, is the specific gravity. It is the ratio of the density of oil to the density of water and is dependent on two temperatures, those at which the densities of the oil sample and the water are measured. When the water temperature is 4 °C (39 °F), the specific gravity is equal to the density in the cgs system, because the volume of 1 g of water at that temperature is, by definition, 1mL. Thus, the density of water, for example, varies with temperature, whereas its specific gravity is always unity at equal temperatures. The standard temperatures for specific gravity in the petroleum industry in North America are 60/60 °F and 15.6/15.6 °C. Although density and specific gravity are used extensively in the oil industry, the API gravity is considered the preferred property (Howard, 2003).

Carbon Residue:

Carbon residue is the percentage of carbon by weight for coke, asphalt, and heavy fuels found by evaporating oil to dryness under standard laboratory conditions. Carbon residue is generally referred to as CCR (Conradson carbon residue). It is a rough indication of the asphaltic compounds and the materials that do not evaporate under conditions of the test, such as metals and silicon oxides (Howard, 2003).

Viscosity:

The viscosity is the measure of the resistance of a liquid to flow, hence indicating the “pumpability” of oil (Howard, 2003).

Pour Point:

This is defined as the lowest temperature (5 F) at which the oil will flow. The lower the pour point, the lower the paraffin content of the oil.

Metals:

In particular, arsenic, nickel, lead, and vanadium are potential poisons for process catalysts. Metal contents are reported in parts per million (Howard, 2003).

Nitrogen:

It is the weight of total nitrogen determined in a liquid hydrocarbon sample (in ppm). Nitrogen compounds contribute negatively to process catalysts (Howard, 2003).

Salt Content:

Salt content is typically expressed as pounds of salt (sodium chloride, NaCl) per 1000 barrels of oil (PTB). Salts in crude oil and in heavier products may create serious corrosion problems, especially in the top tower zone and the overhead condensers in distillation columns (Howard, 2003).

Sulphur:

This is the percentage by weight (or ppm) of total sulphur content determined experimentally in a sample of oil or its product. The sulphur content of crude oils is taken into consideration in addition to the API gravity in determining their commercial values. It has been reported that heavier crude oils may have high sulphur content.

Internationally, crude oils are described according to their sulphur contents. This greatly affects their market values. Table below shows the crude types and their sulphur contents. High sulphur crudes or even sour gases are found mostly in dolomite-anhydrite reservoirs i.e. carbonate-sulfate reservoirs. Such reservoirs are common in the Middle East, Venezuela etc. Arabian Heavy crude from Iran contains 2.8-4.9 % sulphur while Venezuelan crude contains up to 5.4 % at times.

Sulphur in crude is an important source of industrial sulphur as it is a by-product of the refining of heavy crudes. Sulphur however generally a degrading factor to quality rating of crudes for the following reasons:

Sulphur oxidises to sulphur dioxides (SO_2) - a gas, which easily hydrolyses to H_2SO_4 (acid) (Howard, 2003).

The sulphur content of crude oil varies inversely as API gravity. This is the reason why low sulphur-high API crudes like Bonny Light and North Sea Brent are widely sought after in the International market (Howard, 2003).

2.3 Crude Oil Comparison and Crude Oil Assay

In order to establish a basis for the comparison between different types of crude oil, it is necessary to produce experimental data in the form of what is known as an "assay." Crude assays are the systematic compilation of data for the physical properties of the crude and its fractions, as well as the yield.

In other words, a crude assay involves the determination of the following:

1. The properties of crude oil
2. The fractions obtained:
 - (a) their percentage yield and
 - (b) properties

Analytical testing only without carrying out distillation may be considered an assay. However, the most common assay is a comprehensive one that involves all of the above-stated parameters. The basis of the assay is the distillation of a crude oil under specified conditions in a batch laboratory distillation column, operated at high efficiency (column with 14 plates and reflux ratio). Pressure in column is reduced in stages to avoid thermal degradation of high boiling components (Howard, 2003).

2.4 Crude Oil Classifications and Characterization

2.4.1 Crude Oil Classifications

Although there is no specific method for classifying crude oils, it would be useful to establish simple criteria to quantify the quality of crude. Numerous attempts have been made to devise a system to classify crude oils into types based on the predominant hydrocarbon series present in the crude. Such attempts have only partially succeeded. In the United States, crude oils are classified into three types:

1. **Paraffinic:** paraffinic hydrocarbons with a relatively lower percentage of aromatics and naphthenes
2. **Naphthenic:** cycloparaffins in a higher ratio and a higher amount of asphalt than in paraffinic crudes
3. **Asphaltic:** fused aromatic compounds and asphalt in higher Amounts

Another method of classification is the following:

1. Paraffinic base
2. Mixed base
3. Naphthenic base

Based on this classification, a rating for the processing of crude oils is envisaged as follows for the production of certain products and their treatment (Douglas, 2008).

2.4.2 Characterization Factors

Correlation indexes or characterization factors are used in the petroleum industry to indicate the crude type or class. There are several correlations between yield and type of crude in terms of aromaticity and parafinicity (Douglas, 2008).

2.5 Distillation Fundamentals

Distillation processes were known from the 2nd millennium BC by Babylonian alchemists, but the first exact description of distillation apparatus was given in the fourth century by Zosimus of Alexandria (Forbes, 1948). In the following centuries the knowledge of the distillation increases and around the eleventh century this technology was used for the first time in the northern Italy to produce alcoholic beverages. The typical application at that time was the production of spirits (Libavius, 1964).

Obviously, today many things are changed in the technology used to perform the separation and distillation process becomes the most used separation technique. Passing from pot stills to the modern distillation columns, the fundamentals of the separation method remain the same: the different chemical substances contained in a mixture are separated according to the difference on their vapour pressure. The vapour pressure is defined as the equilibrium pressure between the molecules leaving and entering the liquid surface at a fixed temperature. This value is related to the substance boiling point and to a high vapour

pressure corresponds to a low boiling point. It is obvious that the vapour pressure and the boiling point depend on the amount of components in the mixture. In this case a benzene-toluene system is considered. The boiling point of benzene is that at which the mole fraction of this compound is 1, the boiling point of toluene can be read at a zero benzene mole fraction. The bubble and the dew curves are indicated. The first includes all the temperatures at which the liquid phase starts to boil; the latter includes the temperature points at which the saturate vapour starts to condense for the considered composition range. The region above the dew point curve is called superheated vapour region, and the region below the bubble point curve is the sub cooled liquid region. Considering a sub cooled liquid mixture (point A), when it is heated its temperature rises without changing the composition until it reaches the bubble point (point B) and starts to boil. The equilibrium composition of the vapour with the boiling liquid corresponds to point C. This diagram can be used to predict the composition of a single equilibrium stage, given a feed composition and the stage temperature (Alphonsus, 1991).

The difference between the vapour and liquid composition is the driving force (D_{ij}) and is the base parameter of distillation processes. This value is inversely proportional to the energy consumption and directly proportional to separability.

$$D_{ij} = y_i - x_i \dots\dots\dots (2.1)$$

The driving force value is the basis of a class of design methods called “driving force based design” for the optimal selection, with respect to the cost of the operation, of the design parameters like the feed location and the corresponding reflux ratio (Bek-Pedersen, 2004).

Another way to express the driving force for the distillation process is by referring to the relative volatility; this parameter is very common in the distillation design methods and its value is proportional to the easy of the separation. The relative volatility of component I with respect to component n is defined as:

$$\alpha_{in} = \frac{\left(\frac{y_i}{x_i}\right)}{\left(\frac{y_n}{x_n}\right)} = \frac{k_i}{k_n} \dots\dots\dots (2.2)$$

If this value is close to 1 it means that the components have a very close vapour pressure, thus they have similar boiling points and their separation by distillation will be difficult.

The relative volatility value can be used to predict the vapour-liquid equilibrium for an ideal mixture. With simple analytical substitutions it is possible to obtain the following equation that it is useful to construct the x-y diagram (at a fixed pressure) that is the base for graphical design methods, like the McCabe-Thiele one, that allows to evaluate the number of theoretical stages necessary for the separation and consequently the height of the column.

$$y_i = \frac{x_i \alpha_{in}}{1 + x_i(\alpha_{in} - 1)} \dots\dots\dots (2.3)$$

Equation 2.3 can be combined with Equation 2.1 to obtain the following general equilibrium condition.

$$D_{ij} = \frac{x_i \beta_{in}}{1 + x_i(\beta_{in} - 1)} \dots\dots\dots (2.4)$$

$$\beta_{in} = f(T, P, x, y, \phi) \dots\dots\dots (2.5)$$

The concept of equilibrium or theoretical stage is of fundamental importance in distillation design. A theoretical stage is defined as a contacting stage in which equilibrium is attained between the two phases involved (Bek-Pedersen, 2004). Thermodynamic correlations can be used to determinate the temperature and the concentrations of the equilibrium streams at a given pressure. Utilizing a single equilibrium stage it is possible to achieve a limited degree of separation, because a net transfer between two phases can occurs only when there is a driving force, that, in this case, is a difference in concentration values. When the equilibrium is reached the driving force and consequently the net mass transfer is equal to zero. Anyway it is possible to repeat the process by taking the vapour from the separation stage and by feeding it to another separation stage where a partial condensation takes place. The liquid obtained can be fed to another separation stage to be partially vaporized. By repeating the procedure, a cascade of equilibrium stages is obtained with the vapour phase continuously enriched in the most volatile component and the liquid phase with the less volatile component becoming more concentrated. In principle, by creating a large enough cascades an almost complete separation can be carried out (Bek-Pedersen, 2004).

2.5.1 Distillation Process Classification

There are a lot of classifications about the distillation process in function of which aspect is taken into account. One of the most common classifications is according to the nature of the feed processed. We refer to binary distillation if there are only two components to be separated, this is the simplest case and a lot of theoretical works refer to the binary case. In the multi-component distillation process the feed mixture contains more than two components; a classical example is the oil processing that contains thousand of components. The behaviour of the feed components determines successive distinction. It is common to distinguish between ordinary distillation and azeotropic distillation. The first refers to components with a regular thermodynamic behaviour, the latter regards components for which the liquid and the vapour phases can reach the same composition at a fixed point in the distillation process. No further separation can occur after this point without changing the operative variables of the process.

Another classification is about the way by which the feed and the products are introduced and withdrawn in the process. In continuous distillation processes the feed/s and the products are going into and out at the same time. Instead in batch distillation the feed and the withdrawn are not supposed at the same time. Both distillation processes are industrially used for different purposes. Usually continuous processes are more convenient for large feed flow rates, batch processes are preferred for low productions or for plants that, with the same apparatus, must produce different substances. Another technical classification can be made about the type of column internals. The separation principle of distillation processes is the mass transfer across the vapour-liquid interface, for this reason any distillation equipment has to provide a high mass transfer rate. Tray columns and packed columns are most often used for distillation. Proper choice of column internals is very important for the effective and economical separation, anyway is not always possible to define an unambiguous choice because both tray and packed columns can be modified to fulfil different requirements. Useful guidelines on this topic are reported in the literature (Stichlmair, 1998).

2.6 Refining Units

2.6.1 Crude Distillation Unit

The Crude Distillation Unit (CDU) is the first separation process that takes place in a refinery plant.

Figure 2.1 shows a simplified view of the plant. A 42 °API crude, stored at a temperature of about 50 °C, is heated in the first section of the heat exchanger network that utilizes as heating stream the lightest stream from the main column; in this way the crude oil reaches a temperature of about 120 °C and is fed to the desalter to remove inorganic salts, impurities and soluble metals. Then the desalted crude flows through the second section of the heat exchanger network. Due to the great attention on energy integration, by maximising heat exchanges between the crude oil and the product streams from the main column, the crude can reach a maximum temperature of about 240 °C. This temperature is still too low to achieve the grade of crude vaporization necessary for the separation in the main column and thus a furnace is always necessary. The temperature of the exiting stream from the furnace is about 345 °C and fuel oil or fuel gas, depending on the refinery availability, is used as energy source. All the heat needed for the separation is given in the furnace, so no reboilers occur in the main column (Heather, 2009).

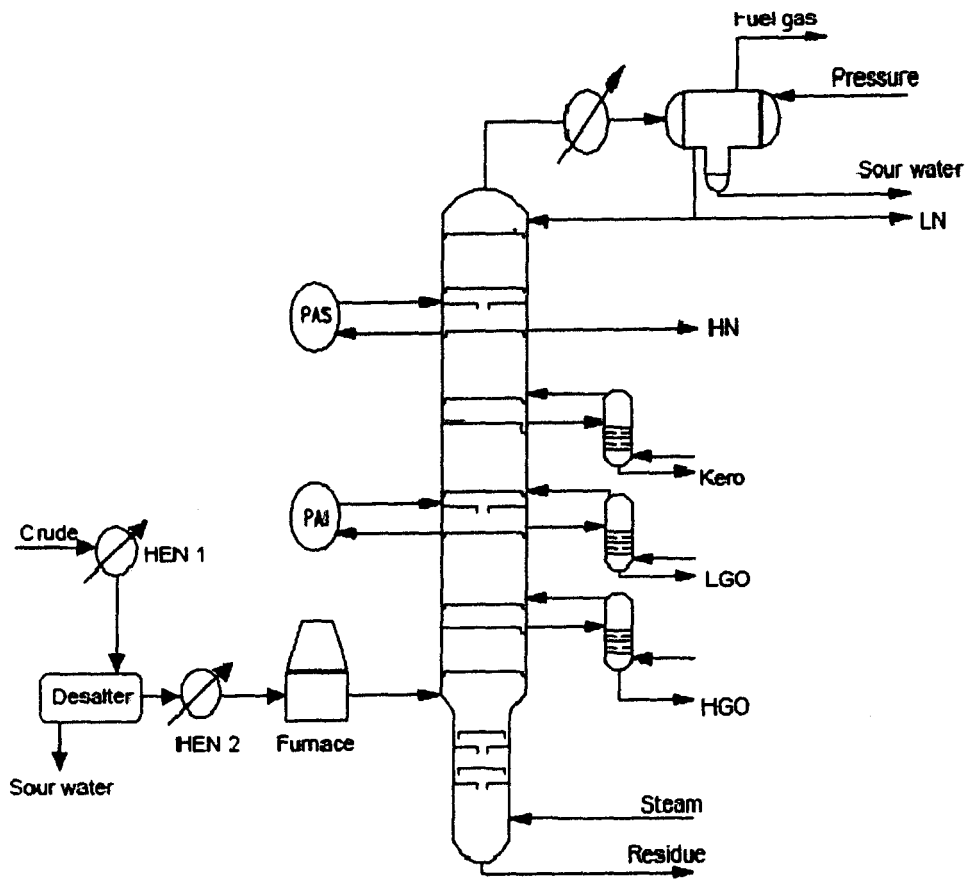


Fig. 2.1: Crude Distillation Unit configuration.

2.6.2 Crude Distillation Unit with a Preflash Device

It was evaluated that the energy requirement for crude distillation unit plants is an amount of fuel equivalent to the 2 % of the total crude processed (Bagajewicz, 2001). For this reason there is a continuous interest to identify ways to improve the energy efficiency of the existing plants.

Different type of solutions were proposed during the years; in the first half of 80's the most popular strategy to increase energy recovery between process fluids was the Pinch method, after that other solutions were considered including also the modification of the distillation design. One popular revamping solution is the employing of preflash device, a drum or a column, (Linnhoff, 1983) to save energy in crude distillation plants following the first indication given by Brugma. The basic idea of a preflash device implementation is to remove the light components of the crude before entering in the furnace. The vapour stream obtained can then be introduced at the furnace outlet or in an appropriate location of the main column. In this way it is possible to reduce the heat duty of the distillation unit and

the furnace and to have also an improvement of the hydraulic performance of the heat exchanger network.

It is a common opinion that the best preflash location is downstream the desalting process in order to remove, with the light components, also the water carryover that can cause corrosion in the following devices or vaporization in the control valves.

Two main approaches for the preflash implementation have been considered in the literature; the first concerns the impact of the preflash device on the heat exchanger network and the second is about the impact of this device on the main column performance. In the former category Harbert (1978), Feintuch *et al* (1985) and Yahyaabadi (2006) made a very clear review of the problem and give useful information to complete the general knowledge about the behaviour of the system. Feintuch *et al.*, consider the modification of the preheating crude network to increase the energy recovery; they focus their attention on the maximum utilization of the existing equipment. In particular they consider the hydraulic limitations and the pressure drop of the modified system and observe that the implementation of a preflash drum just downstream the desalter is able to decrease the operative pressure of the heat exchangers between the flash drum and the furnace. Thus no new equipments are necessary to increase the energy savings in the whole heat exchanger network. They also report that this solution is cost effective with a payout period of less than 3 years. Yahyaabadi studies common problems in preheating trains and the best placement of the preflash drum in the preheat train network below the desalter. He finds that the location of the preflash device has only a small effect on the hot and cold utility consumptions but it is of great importance on the pressure drop and on the average skin temperature of the furnace. There are also some cases in which it is possible to remove part of the heat exchanger network obtaining additional savings on the operating costs. Recently it was also considered the possibility to employ a preflash system for heavy oils, (Waintraub, 2007) that up to now was not taken into account due to the small amount of vaporization that can be achieved. However also in this case it is possible to eliminate water carried over from the desalter and some light components, thus reducing the pressure at the furnace inlet. The second approach considered in the study is the behaviour of the main

column when the preflash device is introduced. In this case there are many criticisms about the possibility to achieve a real energy saving. We refer in particular to the meaningful works of Ji and Bagajewiz and of Golden. The former work includes the preflash drum or the preflash column in a design method for the whole system including the main column. They make a detailed analysis explaining the effect of the light compounds of the crude, called carrier-effect, in improving the separation of the gasoil fraction and also compared different carrier gases to improve the gasoil yield. In another work, Ji (2002), the same authors consider the preflash and the main column system integrated with the vacuum column, and find that the whole system has an energy request slightly smaller than the base design without the preflash system. The position of Golden (1997) on the performance of the preflash device is more critical. He analyzes many parameters that influence the performance of the main column, like the flash drum temperature, the flashed vapour feed location, the effect of flashed crude entrainment in the vapour stream and the quench effect of the flashed vapour in the main column for a fixed outlet furnace temperature. He made a complete study of the preflash drum theory and reports a revamp case. This case study fails due to a feed lighter than the design case highlighting the necessity to design the preflash system for the light oil processed. Anyway this result cannot be considered meaningful of a poor preflash performance. In fact every device has a maximum efficiency in the design operative range, so it is usual that poor performance happens in unexpected situations. The study presented in this thesis starts by considering a real plant crude unit with high energy consumption due to the high furnace duty, and the possibility of energy savings utilizing a preflash device is evaluated. This problem differs from the previous works already published, because utilizes real data plant and describes how it is possible to obtain a compromise between production and energy savings without changing the main column layout.

2.6.3 Preflash Column

The preflash column, differently from the preflash drum, realizes a real separation and it is possible to set the cut point of the desired product. There are some reported cases of refineries which use a preflash column in their plant or consider this device in a revamping

project to unload the atmospheric furnace, to eliminate vaporization at the furnace inlet control valves, to increase the naphtha production and to debottleneck the crude column overhead system (Martin, 1999 and Barletta, 2002). The main problems of the preflash column are related to the presence of only a few plates between the flash zone and the withdrawn and the high reflux ratio flow rate (Sloley, 2002). Another aspect to consider is the naphtha reduction in the top of the main column. So, if we want to keep the same end point for the naphtha stream, the top temperature decreases with possible condensation phenomena and consequent corrosion possibility. Usually it is better to assure a sufficient column top temperature to avoid corrosion and a long time running apparatus (Sloley, 2002).

2.6.4 Two-phase gas–oil separation

At the high pressure existing at the bottom of the producing well, crude oil contains great quantities of dissolved gases. When crude oil is brought to the surface, it is at a much lower pressure. Consequently, the gases that were dissolved in it at the higher pressure tend to come out from the liquid. Some means must be provided to separate the gas from oil without losing too much oil. In general, well effluents flowing from producing wells come out in two phases: vapor and liquid under a relatively high pressure. The fluid emerges as a mixture of crude oil and gas that is partly free and partly in solution. Fluid pressure should be lowered and its velocity should be reduced in order to separate the oil and obtain it in a stable form. This is usually done by admitting the well fluid into a gas–oil separator plant (GOSP) through which the pressure of the gas–oil mixture is successively reduced to atmospheric pressure in a few stages. Upon decreasing the pressure in the GOSP, some of the lighter and more valuable hydrocarbon components that belong to oil will be unavoidably lost along with the gas into the vapour phase. This puts the gas–oil separation step as the initial one in the series of field treatment operations of crude oil. Here, the primary objective is to allow most of the gas to free itself from these valuable hydrocarbons, hence increasing the recovery of crude oil. Crude oil as produced at the wellhead varies considerably from field to field due not only to its physical characteristics but also to the amount of gas and salt water it contains. In some fields, no salt water will

flow into the well from the reservoir along with the produced oil. This is the case we are considering in this chapter, where it is only necessary to separate the gas from the oil; (i.e., two-phase separation) (Marcel, 2003).

When, on the other hand, salt water is produced with the oil, it is then essential to use three-phase separators—a case studied in Chapter 4. Accordingly, oil-field separators can be classified into two types based on the number of phases to separate:

1. Two-phase separators, which are used to separate gas from oil in oil fields, or gas from water for gas fields.
2. Three-phase separators, which are used to separate the gas from the liquid phase, and water from oil.

Oil from each producing well is conveyed from the wellhead to a gathering center through a flow line. The gathering center, usually located in some central location within the field, will handle the production from several wells in order to process the produced oil–gas mixture. Separation of the oil phase and the gas phase enables the handling, metering, and processing of each phase independently, hence producing marketable products (Marcel, 2003).

2.6.5 Theory of Gas–Oil Separation

In order to understand the theory underlying the separation of well effluent hydrocarbon mixtures into a gas stream and oil product, it is assumed that such mixtures contain essentially three main groups of hydrocarbon, as illustrated in Figure 6:

1. Light group, which consists of CH_4 (methane) and C_2H_6 (ethane)
2. Intermediate group, which consists of two subgroups: the propane/butane ($\text{C}_3\text{H}_8/\text{C}_4\text{H}_{10}$) group and the pentane/hexane ($\text{C}_5\text{H}_{12}/\text{C}_6\text{H}_{14}$) group.
3. Heavy group, which is the bulk of crude oil and is identified as C_7H_{16} . In carrying out the gas–oil separation process, the main target is to try to achieve the following objectives:
 1. Separate the C1 and C2 light gases from oil
 2. Maximize the recovery of heavy components of the intermediate group in crude oil
 3. Save the heavy group components in liquid product

To accomplish these objectives, some hydrocarbons of the intermediate group are unavoidably lost in the gas stream. In order to minimize this loss and maximize liquid recovery, two methods for the mechanics of separation are compared:

1. Differential or enhanced separation

2. Flash or equilibrium separation

In differential separation, light gases (light group) are gradually and almost completely separated from oil in a series of stages, as the total pressure on the well-effluent mixture is reduced. Differential separation is characterized by the fact that light gases are separated as soon as they are liberated (due to reduction in pressure). In other words, light components do not come into contact with heavier hydrocarbons; instead, they find their way out. For flash separation, on the other hand, gases liberated from the oil are kept in intimate contact with the liquid phase. As a result, thermodynamic equilibrium is established between the two phases and separation takes place at the required pressure. Comparing the two methods, one finds that in differential separation, the yield of heavy hydrocarbons (intermediate and heavy groups) is maximized and oil-volume shrinkage experienced by crude oil in the storage tank is minimized. This could be explained by the fact that separation of most of the light gases takes place at the earlier high-pressure stages; hence, the opportunity of losing heavy components with the light gases in low-pressure stages is greatly minimized. As a result, it may be concluded that flash separation is inferior to differential separation because the former experiences greater losses of heavy hydrocarbons that are carried away with the light gases due to equilibrium conditions. Nevertheless, commercial separation based on the differential concept is very costly and is not a practical approach because of the many stages required. This would rule out differential separation, leaving the flash process as the only viable scheme to affect gas-oil separation using a small number of stages. A close approach to differential separation is reached by using four to five flash separation stages. A comparison between the mechanisms of separation by the two methods is schematically simplified and presented (Marcel, 2003).

2.6.6 Methods Of Separation

The traditional process for recovering crude oil from high-pressure well streams is based on the flash separation concept explained in Section 3.3, which consists of a series of flash separators operating over a pressure range from wellhead pressure to atmospheric pressure. However, with the increased desirability of recovering natural gas and natural gas liquids (NGLs), other methods have been proposed as modification to the basic flash separation technique. Separation methods could be broadly classified as follows:

1. Conventional methods

2. Modified methods

(a) Adding vapour recompression unit to the conventional methods

(b) Replacing the conventional methods by a stabilizer and a recompression unit the conventional method is a multistage flash separation system and is recommended for comparatively high-pressure fluids. Several stages operated at successively lower pressures affect the separation of oil from gas, thus increasing the oil recovery. In general, the number of stages in a multistage conventional separation process is a function of the following:

1. The API gravity of the oil

2. The gas–oil ratio (GOR)

3. The flowing pressure consequently, high-API-gravity oils with high GOR flowing under high pressure would require the greatest number of stages, from three to four (Marcel, 2003).

The first modified method of separation implies adding several stages of gas compression to recompress the separated gas from each flash stage. Liquids from inter-stage vessels between the compressors can be collected and processed as liquid natural gas (LNG) stock. Natural gas will be delivered at the desired pressure depending on its usage. The second modified method of separation is different in concept from the conventional (flash) separation. It makes use of crude stabilizer columns. Normally, these columns have top-feed trays with no rectifying section and no condenser, but are provided with inter-stage reboilers and feed preheaters. Crude stabilization systems are advantageous as GOSPs if

space is critical, as may be encountered on an offshore platform, because they occupy less space than conventional GOSPs.

2.6.7 Types of Gas-Oil Separator

2.6.7.1 Commercial types of Gas–Oil Separator

Based on the configuration, the most common types of separator are horizontal, vertical, and spherical. A concise comparison among these three types is presented in Table 3. Large horizontal gas–oil separators are used almost exclusively in processing well fluids in the Middle East, where the gas–oil ratio of the producing fields is high. Multistage GOSPs normally consists of three or more separators.

The following is a brief description of some separators for some specific applications. In addition, the features of what is known as “modern” GOSP are highlighted. Test Separators These units are used to separate and measure at the same time the well fluids. Potential test is one of the recognized tests for measuring the quantity of both oil and gas produced by the well in 24 hours period under steady state of operating conditions. The oil produced is measured by a flow meter (normally a turbine meter) at the separator’s liquid outlet and the cumulative oil production is measured in the receiving tanks.

An orifice meter at the separator’s gas outlet measures the produced gas, physical properties of the oil and GOR are also determined (Heather, 2009).

2.6.7.2 Low-Temperature Separators

Low-temperature separators (LTSs) are used to effectively remove light condensable hydrocarbons from a high-pressure gas stream (gas condensate feed). Liquid (condensate) separation is made possible by cooling the gas stream before separation. Temperature reduction is obtained by what is known as the Joule–Thomson effect of expanding the well fluid as it flows through the pressure-reducing choke or valve into the separator. Condensation of the vapours takes place accordingly, where the temperature is in the range 0–10 °F (Heather, 2009).

2.6.7.3 Modern GOSPs

Safe and environmentally acceptable handling of crude oils is assured by treating the produced crude in the GOSP and related crude-processing facilities. The number one

function of the GOSP is to separate the associated gas from oil. As the water content of the produced crude increases, field facilities for control or elimination of water are to be added. This identifies the second function of a GOSP. If the effect of corrosion due to high salt content in the crude is recognized, then modern desalting equipment could be included as a third function in the GOSP design. One has to differentiate between “dry” crude and “wet” crude. The former is produced with no water, whereas the latter comes along with water. The water produced with the crude is a brine solution containing salts (mainly sodium chloride) in varying concentrations (Heather, 2009).

The input of wet crude oil into a modern GOSP consists of the following:

1. Crude oil.
2. Hydrocarbon gases.
3. Free water dispersed in oil as relatively large droplets, which will separate and settle out rapidly when wet crude is retained in the vessel.
4. Emulsified water, dispersed in oil as very small droplets that do not settle out with time. Each of these droplets is surrounded by a thin film and held in suspension.

5. Salts dissolved in both free water and in emulsified water. The functions of a modern GOSP could be summarized as follows:

1. Separate the hydrocarbon gases from crude oil
2. Remove water from crude oil
3. Reduce the salt content to the acceptable level (basic sediments and water)

It should be pointed out that some GOSPs do have gas compression and refrigeration facilities to treat the gas before sending it to gas processing plants. In general, a GOSP can function according to one of the following process operation:

1. Three-phase, gas–oil–water separation
2. Two-phase, gas–oil separation
3. Two-phase, oil–water separation
4. Demulsification
5. Washing
6. Electrostatic coalescence

2.7 Process Furnaces

Process furnaces are equipments where fluids are heated by the gases produced by combustion of liquid or gaseous fuel. They are termed direct fired furnaces, as the heat of the combustion gases is directly transmitted to the cold fluid that circulates in a tubular coil. These furnaces are different from indirectly fired furnaces where the process fluid, which is itself heated directly by the fuel combustion gases. Direct fired furnaces must also be distinguished from mixing type furnace (such as catalyst regenerators) where the process fluids are raised to the specified temperature by mixing with the fuel combustion gases (Bonnet, 1982).

The heat transmitted to the heated fluid can be used:

- Simply to heat it: the temperature of the process fluid is raised so that it can undergo a physical or chemical change later on (steam super-heater, catalytic reformer gas heater, liquid hydrocarbon heater feed for the catalytic unit, heat transfer fluid, etc)
- To accomplish a chemical change: by pyrolysis (steam cracking furnaces to manufacture ethylene, dichloroethane and acetic acid cracking furnace, etc).

In other cases, the change is accomplished in the presence of a catalyst production of synthesis gas from natural gas or naphtha, isopropyl alcohol cracking furnaces, etc). In the most common instance, several operations are carried out either simultaneously or successively in the same furnace, for example: visbreaking furnace where sensible, reaction (thermal cracking) and vaporization heats are supplied.

Several different fluids can also be heated simultaneously in the same furnace which is said to be multi-service. However, here only one of the services can generally be controlled, with heat automatically supplied to the others. An accessory device is then incorporated to control their outlet temperature (Bonnet, 1982).

2.7.1 Description of a Furnace

Generally speaking, furnaces comprise the following main parts:

- a. A radiation section consisting essentially of a combustion chamber where the tubes are located. The tubes are separate from each other but connected by elbows. The process fluid circulates inside this tube bundle and heat is transferred from flue

gases and the tubes. The temperature of the flue gases exiting the radiation section is high (700 to 1,100 °C) and so the efficiency is low (approximately 50 % of the low heating value) (Bonnet, 1982).

- b. In order to recover the sensible heat from the flue gases, they circulate at high speed (parallel or perpendicular) through a tube bundle where heat is transferred chiefly by convection. This sector is then called the convection section. The tubes may be lined or studded in order to increase the heat transfer surface area on the combustion gas side, the flue gases usually being the fluid with the greater thermal resistance. The efficiency of a furnace with a convection section is obviously higher than that of a furnace radiation section alone. Efficiency depends on the inlet temperature of the heated fluid (flue gases cannot be cooled below the temperature of cold fluid), but also on the size of the heat exchanger surface area in the furnace (Bonnet, 1982).
- c. A stack to discharge the flue gases (Bonnet, 1982).

2.7.2 Types of Furnaces

There are numerous layouts for tubes in the radiation and convection sections and for one section in relation to the other. There are consequently many types of furnaces, that can however be classified into the following different categories:

- **Vertical cylinder furnaces:** the radiation section is in the shape of a cylinder with a vertical axis. The burners are located on the floor at the base of the cylinder. The heat exchange area covers the vertical walls and therefore exhibits circular symmetry with respect to the heating assembly.
- **Box type furnaces with vertical tube:** here the general shape of the radiation section is a parallelepiped. The burners are situated on the floor and the heat exchange area covers the vertical side walls. Sometimes the combustion chamber is divided up into several cells by rows of tubes parallel to one of the side wall faces.
- **Cabin furnaces with horizontal tubes:** here the general shape of the radiation section is a parallelepiped, with the greatest length being horizontal and located on the floor,

on the lower part of the longest side walls or on the end walls where there are no tubes.

- **Radiant wall furnaces:** the heat exchange surface area is located on the median plane of the combustion chamber. The burners are distributed on the longitudinal sidewalls.
- **Double fired furnaces:** here the tube plane constituting the heat exchange surface area is in the median plane of the rows of burners (Bonnet, 1982).

Chapter Three

3.0 MATERIAL BALANCE

The general conservation equation for any process system can be written as:

$$\text{Input} + \text{Generation} = \text{Output} + \text{Accumulation} + \text{Consumption}.$$

For a steady-state process, the accumulation term will be zero except in nuclear processes where mass is neither generated nor consumed; but if a chemical reaction takes place a particular chemical species may be formed or consumed in the process. If there is no chemical reaction the steady-state balance reduces to

$$\text{Input} = \text{Output} \text{ (Sinnott, 2003).}$$

It must be noted that the entire operation of this design has been assumed to be at steady state since there is no chemical reaction taking place in any unit.

3.1 Crude Oil Assay

Table 3.1: Properties of the Crude oil

API gravity	49°
Viscosity kg/ms	-
Mass Enthalpy kJ/kg	-1549
Mass Heat Capacity kJ/kg-C	2.952

Source: (NNPC, 2010)

Table 3.2: Cut distributions

Component	Mass Fraction
Off gas	8.313×10^{-9}
Waste water	6.058×10^{-3}
Gasoline	1.761×10^{-1}
Naphtha	9.353×10^{-2}
Kerosene	1.247×10^{-1}
Diesel	1.684×10^{-1}
AGO	5.095×10^{-2}
Residue	3.802×10^{-1}

Source: (NNPC, 2010)

Table 3.3: Preheat crude properties

Property	Value
Temperature	232.2 °C
Pressure	517.1 kPa
Liquid flow rate	200 m ³
Density of Vapour phase	689.4 kg/m ³
Density of Liquid phase	788.4 kg/m ³

Source: (NNPC, 2010)

3.2 Material Balance around Preflash

From Table 3.1, the API gravity value is related to specific gravity as follows:

$$\text{API gravity} = \frac{141.5}{\text{Specific gravity of crude}} - 131.5 \dots \dots \dots (3.1)$$

$$\Rightarrow S.G = \frac{141.5}{49.08+131.5} = 0.7836$$

$$\therefore \text{Density of Crude} = 0.7836 \times 1000 = 783.6 \text{ kg/m}^3$$

Basis: 200 m³/h

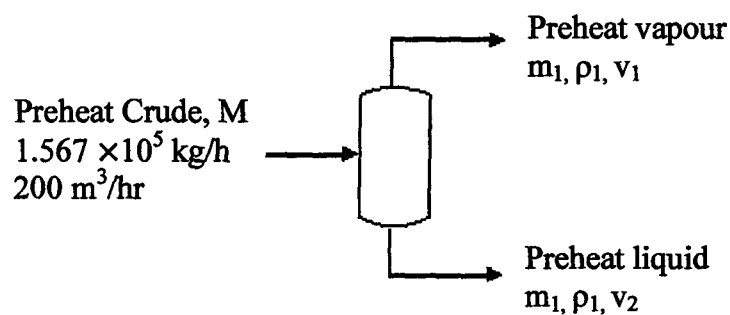
The basis above cannot be used for the material balances without converting it to mass flow rate; this is because of the cut distributions of the various products in mass fraction. From the relation of density, ρ ; mass, M and volume, V; we know that:

$$\rho = M / V; \text{ it implies that } M = \rho V.$$

From Table 3.3, the density of the preheat crude is given as 783.6 kg/m³; therefore, taking a liquid flow rate of 200 m³, we have:

$$M = 200 \text{ m}^3/\text{h} \times 783.6 \text{ kg/m}^3 = 1.567 \times 10^5 \text{ kg/h}$$

$$\therefore \text{Basis} = 1.567 \times 10^5 \text{ kg/h}$$



At steady state, $M = m_1 + m_2$

But $m_1 = \rho_1 v_1$ and $m_2 = \rho_2 v_2$

Where m_1 and m_2 are the mass flow rates of the preheat vapour and the preheat liquid respectively; ρ_1 and ρ_2 are their densities and v_1 and v_2 are their respective volumes.

Also, $V = v_1 + v_2$

$$\therefore 689.4v_1 + 788.4v_2 = 1.567 \times 10^5 \dots\dots\dots (3.2)$$

$$v_1 + v_2 = 200 \dots\dots\dots (3.3)$$

Solving equations (1) and (2) simultaneously gives $v_1 = 9.9 \text{ m}^3/\text{hr}$ and $v_2 = 190.1 \text{ m}^3/\text{hr}$

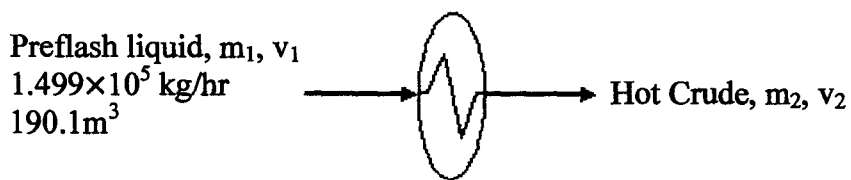
$$\therefore m_2 = 788.4 \times 190.1 = 1.499 \times 10^5 \text{ kg/hr}$$

And $m_1 = M - m_2 = 6825 \text{ kg/hr}$

3.2.1 Summary of Material balance around the Preflash

	Feed	Mass flow (kg/hr)	Volume flow (m ³ /hr)
Inlet stream	Preheat crude	1.567e5	200
Outlet streams	Preflash Vapour	6825	9.9
	Preflash Liquid	1.499e5	190.1

3.3 Material Balance around Furnace



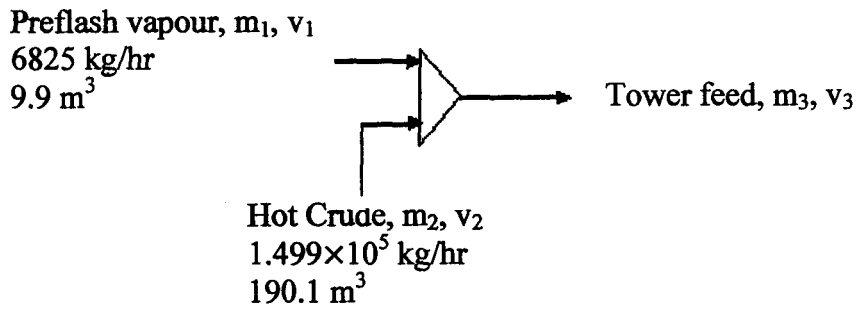
At steady state, $m_1 = m_2$ and $v_1 = v_2$

$$\therefore m_2 = 1.499 \times 10^5 \text{ kg/hr and } v_2 = 190.1 \text{ m}^3$$

3.3.1 Summary of Material balance around the Furnace

	Feed	Mass flow (kg/hr)	Volume flow (m ³ /hr)
Inlet Stream	Preflash Liquid	1.499×10^5	190.1
Outlet Stream	Hot Crude	1.499×10^5	190.1

3.4 Material Balance around Mixer



At steady state condition, $m_1 + m_2 = m_3$ and $v_1 + v_2 = v_3$

$$m_3 = (6825 + 1.499 \times 10^5) \text{ kg/hr} = 1.567 \times 10^5 \text{ kg/hr}$$

$$v_3 = (9.9 + 190.1) \text{ m}^3/\text{hr} = 200 \text{ m}^3/\text{hr}$$

3.4.1 Summary of Material balance around the Mixer

	Feed	Mass flow (kg/hr)	Volume flow (m ³ /hr)
Inlet streams	Preflash Vapour	6825	9.9
	Hot Crude	1.499×10^5	190.1
Outlet stream	Tower Feed	1.567105	200

3.5 Material Balance around Steam Generator

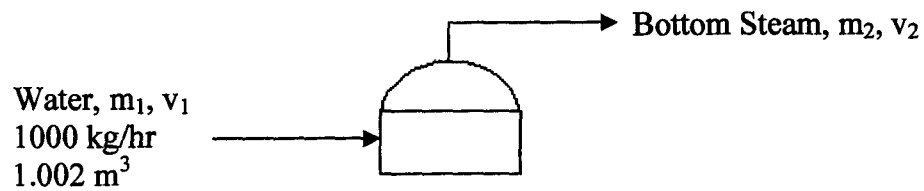
Basis: 1000 kg/hr of water at 70 °C

Density of water at 70 °C, $\rho_w = 998.0 \text{ kg/m}^3$

But $V = M/\rho$ where M and ρ are mass flow rate and density respectively.

$$\therefore V = 1000 \text{ kg/hr} \div 998.0 \text{ kg/m}^3 = 1.002 \text{ m}^3/\text{hr}$$

Basis: 1000 kg/hr (or 1.002 m³/hr)



At steady state, $m_1 = m_2$ and $v_1 = v_2$

$$\therefore m_2 = 1000 \text{ kg/hr and } v_2 = 1.002 \text{ m}^3$$

3.5.1 Summary of Material Balance around the Steam Generator

	Feed	Mass flow (kg/hr)	Volume flow (m ³ /hr)
Inlet Stream	Water	1000	1.002
Outlet Stream	Bottom Steam	1000	1.002

3.6 Material Balance around the Tower (ADU)

The feed that goes into the atmospheric distillation unit (ADU) is a combination of the crude oil and the bottom steam.

Crude oil inlet feed: $1.567 \times 10^5 \text{ kg/hr}$

Steam inlet feed: 1000 kg/hr

Total inlet feed for Tower = $(1.567 \times 10^5 + 1000)$ kg/hr

$\therefore F$ (Tower feed) = 1.577×10^5 kg/hr

$$\Rightarrow F = \sum F_{xf} + B$$

Where:

F = Tower feed

xf = mass fraction of distillates in the feed

Fxf = Distillates

B = Bottom(s)

From the cut distribution in table 3.1b, we have the following results:

$$\text{Off gas} = 8.313 \times 10^{-9} \times 1.577 \times 10^5 = 0.001311 \text{ kg/hr}$$

$$\text{Waste water} = 6.058 \times 10^{-3} \times 1.577 \times 10^5 = 955.3466 \text{ kg/hr}$$

$$\text{Gasoline} = 1.761 \times 10^{-1} \times 1.577 \times 10^5 = 2.777 \times 10^4 \text{ kg/hr}$$

$$\text{Naphtha} = 9.353 \times 10^{-2} \times 1.577 \times 10^5 = 1.475 \times 10^4 \text{ kg/hr}$$

$$\text{Kerosene} = 1.247 \times 10^{-1} \times 1.577 \times 10^5 = 1.967 \times 10^4 \text{ kg/hr}$$

$$\text{Diesel} = 1.684 \times 10^{-1} \times 1.577 \times 10^5 = 2.656 \times 10^4 \text{ kg/hr}$$

$$\text{AGO} = 5.095 \times 10^{-2} \times 1.577 \times 10^5 = 8035 \text{ kg/hr}$$

$$\text{Residue} = 3.802 \times 10^{-1} \times 1.577 \times 10^5 = 5.996 \times 10^4 \text{ kg/hr}$$

$$\text{Total} = 1.577 \times 10^5 \text{ kg/hr}$$

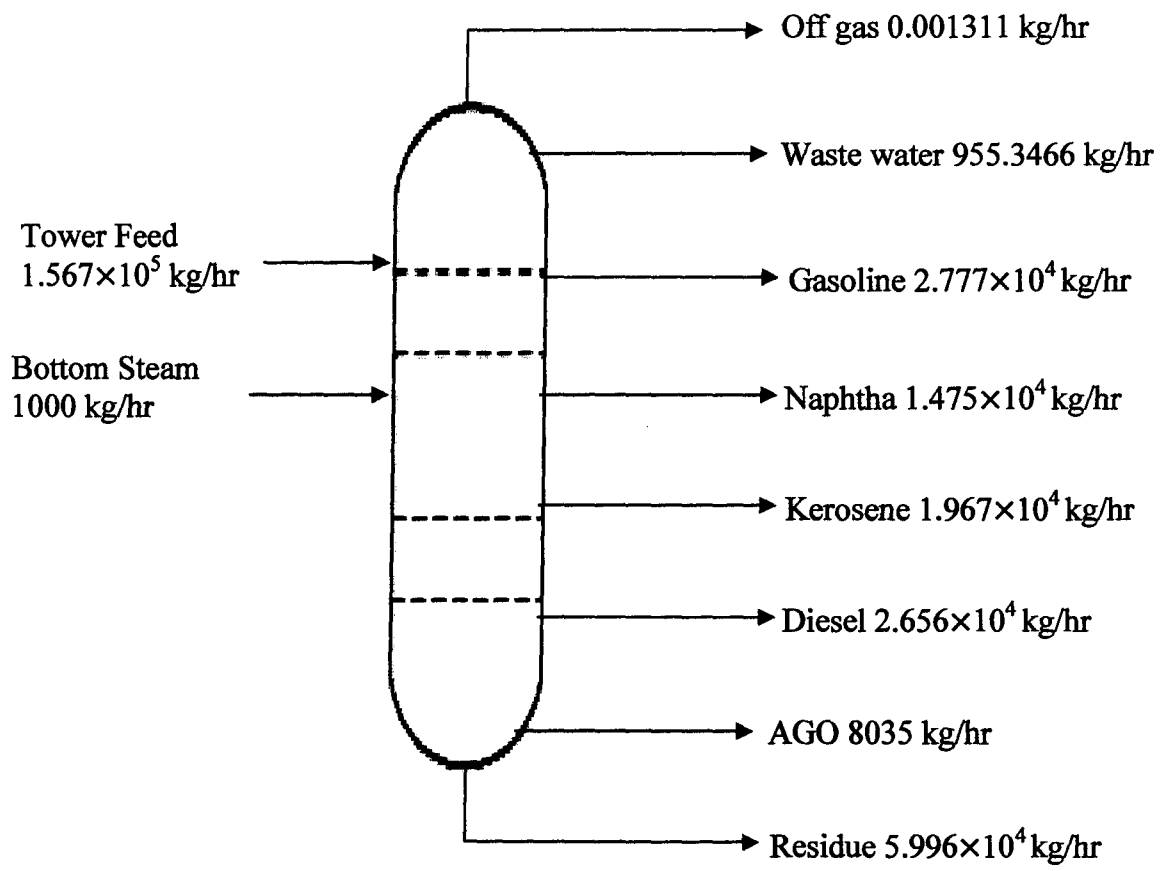


Figure 3.1: Atmospheric Distillation Tower.

Chapter Four

4.0 ENERGY BALANCE

Like material balance, the general conservation equation for any process system (Physical or Chemical) can be written as:

$$\text{Input} + \text{Generation} = \text{Output} + \text{Accumulation} + \text{Consumption.}$$

but for a steady state operation,

$$\text{Input} = \text{Output (Sinnott, 2003).}$$

The amount of energy (in the form of heat) that a given body can absorb at a set temperature depends on the Enthalpy of that body at the set temperature; enthalpy being a function of what is known as the Specific Heat Capacity of the body which is constant for a given body.

$$\text{i.e } H = C_p T$$

Where H = Enthalpy

C_p = Specific Heat Capacity (C_p is constant for all bodies of the same type) and

T = Temperature

4.1 Preheat Crude

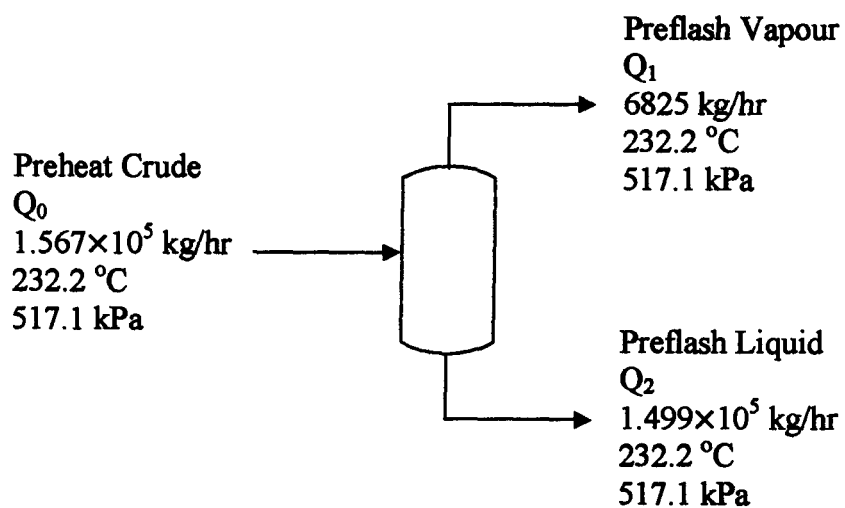
The Table below shows some properties of the preheated crude as collected from the Nigerian National Petroleum Co-operation (NNPC).

Table 4.1: Properties of the Preheat Crude

Internal Energy, kJ	-1.5497×10^3
Pressure, kPa	517.1
Temperature, °C	232.2
Specific Heat Capacity, kJ/kg-°C	10.4540
Bulk Density, kg/m ³	783.6
Mass flow rate, kg/hr	1.567×10^5
Molecular weight	300.0

Source: (NNPC, 2010)

4.2 Energy Balance around the Preflash



Q_0 , Q_1 and Q_2 are the heat flow rates of the preheat crude, the Preflash vapour and the Preflash liquid respectively.

From thermodynamics, Enthalpy, $H = U + Pv$

Where U = Internal energy

P = Pressure

v = Volume per unit mass

From Table 4.1a, we have the internal energy to be -1.5497×10^3 kJ and from the previous chapter, we recall the Volume and Mass bases to be $200 \text{ m}^3/\text{hr}$ and 1.567×10^5 kg/hr respectively.

$$\Rightarrow v = \frac{200}{1.567e5} = 1.2763 \times 10^{-3} \text{ m}^3/\text{kg}$$

$$\therefore H = -1.5497 \times 10^3 + 517.1(1.2763 \times 10^{-3}) = -1549.04 \text{ kJ/kg}$$

Also, from heat capacity, we know that, $Q = MC\Delta\theta$

Where Q = quantity of heat required

C = specific heat capacity

$\Delta\theta$ = change in temperature

but $H = C\Delta\theta$

$\Rightarrow Q = MH$, i.e Energy = Mass \times Enthalpy

$$\therefore \text{Inlet Energy stream, } Q_0 = 1.567 \times 10^5 \times -1549.04 = -2.427 \times 10^8 \text{ kJ/hr}$$

At steady state, $Q_0 = Q_1 + Q_2$

Q_1 and Q_2 are the energy streams of the vapour and liquid phase respectively

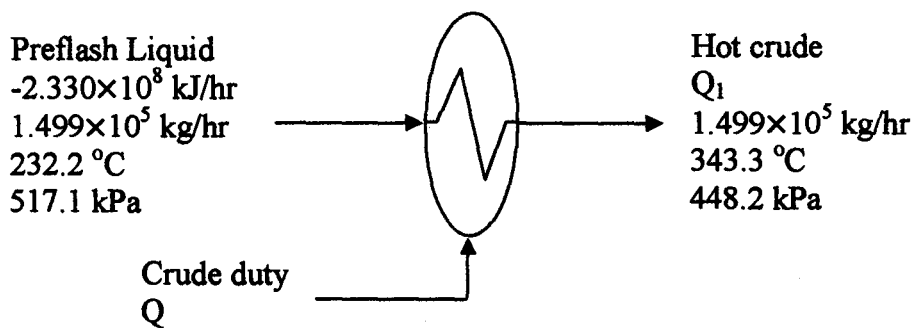
$$Q_1 = 6825 \times -1413.15 = -9.645 \times 10^6 \text{ kJ/hr}$$

$$Q_2 = 1.499 \times 10^5 \times -1554.37 = -2.330 \times 10^8 \text{ kJ/hr}$$

Table 4.2: Summary of Energy and Material balances around the Preflash

	Feed	Heat flow (kJ/hr)	Mass flow (kg/hr)
Inlet stream	Preheat crude	-2.427×10^8	1.567×10^5
Outlet streams	Preflash Vapour	-9.645×10^6	6825
	Preflash Liquid	-2.330×10^8	1.499×10^5

4.3 Energy balance around the Furnace



From the literature review, we have a complete temperature profile of the entire refining process and therefore we know the entry and exit temperature of each stream.

At steady state, $Q - 2.330 \times 10^8 = Q_1$

But $Q_1 = MC\Delta\theta$ and -2.330×10^8 kJ/hr = MH as previously calculated in the Preflash for the Preflash liquid.

$$\therefore MH + Q = MC\Delta\theta$$

Where C is the specific heat capacity of the crude and $\Delta\theta$ is the temperature change.

The crude duty, Q is the amount of heat supplied by the furnace that is required to further heat the crude oil to the desired tower inlet temperature.

MH, from previous section, has been calculated as the energy content of the Preflash liquid, -2.330×10^8 kJ/hr

$$MC\Delta\theta = 1.499 \times 10^5 \times 10.4540 \times (232.2 - 343.3) = -1.741 \times 10^8 \text{ kJ/hr}$$

$$\therefore Q = MC\Delta\theta - MH$$

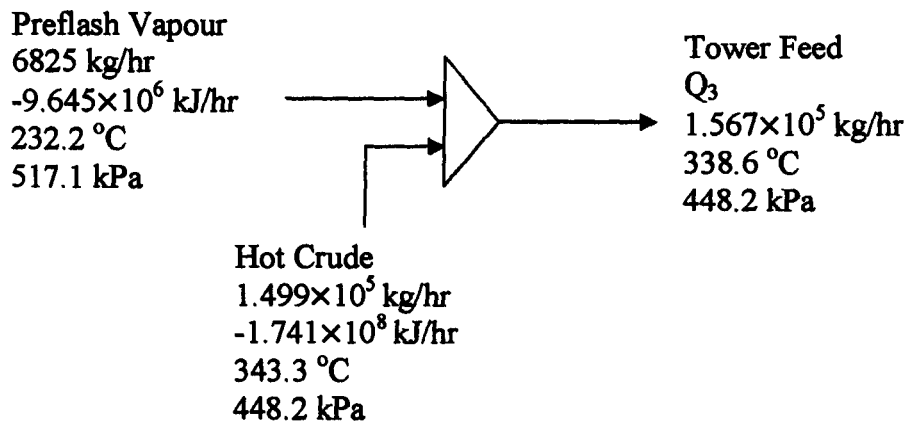
$$= -1.741 \times 10^8 + 2.330 \times 10^8$$

$$= 5.89 \times 10^7 \text{ kJ/hr}$$

Table 4.3: Summary of Energy and Material balances around the Furnace

	Feed	Energy flow (kJ/hr)	Mass flow (kg/hr)
Inlet streams	Preflash Liquid	-2.330×10^8	1.499×10^5
	Crude Duty	5.89×10^7	-
Outlet stream	Hot Crude	-1.741×10^8	1.499×10^5

4.4 Energy Balance around the Mixer



At steady state, $Q_1 + Q_2 = Q_3$

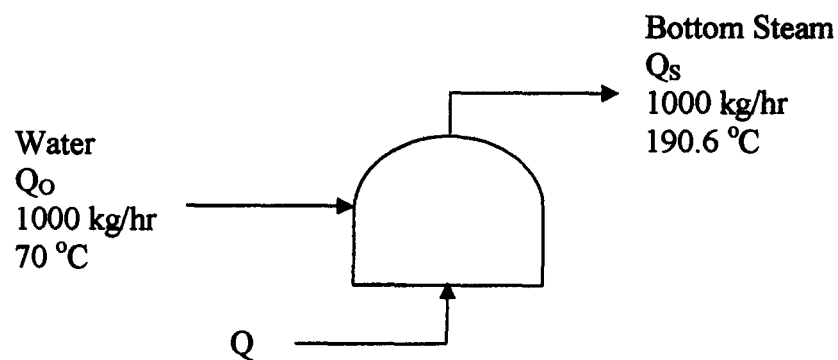
The mixer is not a heating unit, therefore like mass flow rate; the energy streams can be added or subtracted to give a balance as follows

$$Q_3 = -1.741 \times 10^8 + -9.645 \times 10^6 = -1.837 \times 10^8 \text{ kJ/hr}$$

Table 4.4: Summary of Energy and Material balances around the Mixer

	Feed	Energy flow (kJ/hr)	Mass flow (kg/hr)
Inlet streams	Preflash Vapour	-9.645×10^6	6825
	Hot Crude	-1.741×10^8	1.499×10^5
Outlet stream	Tower feed	-1.837×10^8	1.567×10^5

4.5 Energy Balance around the Steam Generator



$$MH_w + Q = MC_w\Delta\theta + MH_s$$

Where MH_w is the inlet heat stream, H_w being the enthalpy of water at 70°C (-15650 kJ/kg); Q is the quantity of heat produced by the steam generator that is required to generate the steam at 190.6°C ; $MC_w\Delta\theta$ is the outlet heat content of the water at 100°C , C_w being the specific heat capacity of water (4.2 kJ/kg- $^\circ\text{C}$); MH_s is the heat required to cause a change of state from 100°C to 190.6°C , H_s being the specific enthalpy of steam

at 190.6 °C. Together, $MC_w\Delta\theta$ and MH_s make up the heat content of the generated steam.

$$\text{i.e } Q_o = MH_w$$

$$Q_s = MC_w\Delta\theta + MH_s$$

$$MH_w = 1000 \times -15650 = -1.565 \times 10^7 \text{ kJ/hr} = Q_o$$

$$MC_w\Delta\theta = 1000 \times 4.2 \times (100 - 70) = 1.26 \times 10^5 \text{ kJ/hr}$$

$$MH_s = 1000 \times -13216 = -1.322 \times 10^7 \text{ kJ/hr}$$

$$\therefore Q_s = -1.309 \times 10^7 \text{ kJ/hr}$$

$$\Rightarrow Q = MC_w\Delta\theta + MH_s - MH_w$$

$$= 1.26 \times 10^5 - 1.322 \times 10^7 + 1.565 \times 10^7$$

$$\therefore Q = 2.56 \times 10^6 \text{ kJ/hr}$$

Table 4.5: Summary of Energy and Material balances around the steam Generator

	Feed	Energy flow (kJ/hr)	Mass flow (kg/hr)
Inlet streams	Water	-1.565×10^7	1000
	Heat	2.56×10^6	-
Outlet stream	Bottom Steam	-1.309×10^7	1000

4.6 Energy Balance around the Tower

From literature, we know the outlet temperature of each product stream; therefore the table below shows the mass flow rate, the outlet temperature and the enthalpy of each product at their exit temperature.

Table 4.6: Properties of Cut distributions

Feed	Mass Flow rate, kg/hr	Exit Temperature, °C	Enthalpy, kJ/kg
Off Gas	0.001311	73.25	-2615
Waste water	955.3466	73.25	-1.563×10 ⁴
Gasoline	2.777×10 ⁴	73.25	-2011
Naphtha	1.475×10 ⁴	225.3	-1595
Kerosene	1.967×10 ⁴	270.8	-1450
Diesel	2.656×10 ⁴	364.5	-1133
AGO	8035	472.8	-732.5
Residue	5.996×10 ⁴	414.1	-973.6

The total inlet energy is the summation of the tower feed energy and the energy of the bottom steam.

$$\therefore \text{Total inlet energy} = (-1.837 \times 10^8 - 1.309 \times 10^7) \text{ kJ/hr} = -1.968 \times 10^8 \text{ kJ/hr}$$

$$\Rightarrow Q_i = M_i H_i$$

Where M_i is the mass flow rate of the individual products and H_i is their respective enthalpy.

$$\therefore Q_{\text{Off gas}} = 0.001311 \times -2615 = -3.428 \text{ kJ/hr}$$

$$Q_{\text{Waste water}} = 955.3466 \times -1.563 \times 10^4 = -1.49310^7 \text{ kJ/hr}$$

$$Q_{\text{Gasoline}} = 2.777 \times 10^4 \times -2011 = -5.585107 \text{ kJ/hr}$$

$$Q_{\text{Naphtha}} = 1.475 \times 10^4 \times -1595 = -2.353107 \text{ kJ/hr}$$

$$Q_{\text{Kerosene}} = 1.967 \times 10^4 \times -1450 = -2.85210^7 \text{ kJ/hr}$$

$$Q_{\text{Diesel}} = 2.656 \times 10^4 \times -1133 = 3.010^7 \text{ kJ/hr}$$

$$Q_{\text{AGO}} = 8035 \times -732.5 = -5.88610^6 \text{ kJ/hr}$$

$$Q_{\text{Residue}} = 5.996 \times 10^4 \times -973.6 = -5.83810^7 \text{ kJ/hr}$$

Chapter Five

5.0 PROCESS FLOW DIAGRAM

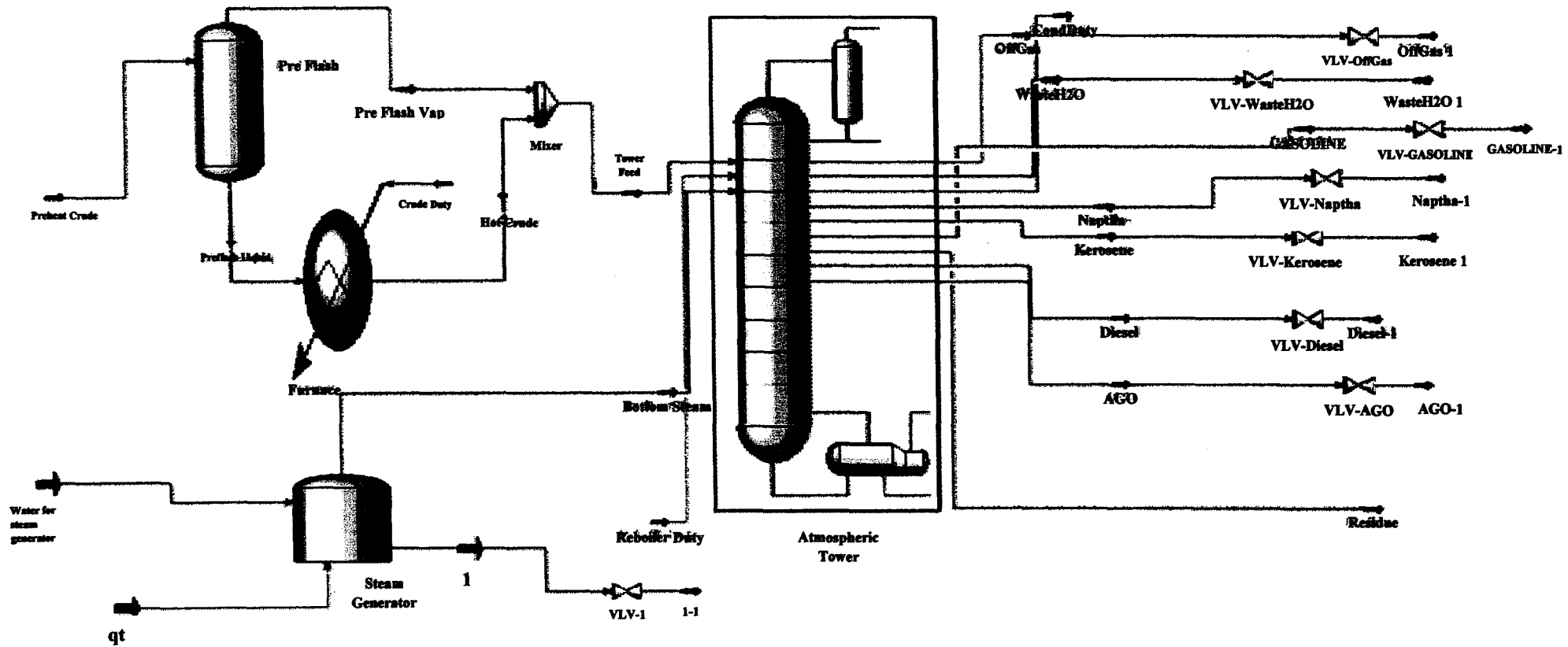


Fig 5.1: Process Flow Diagram for Feed Streams

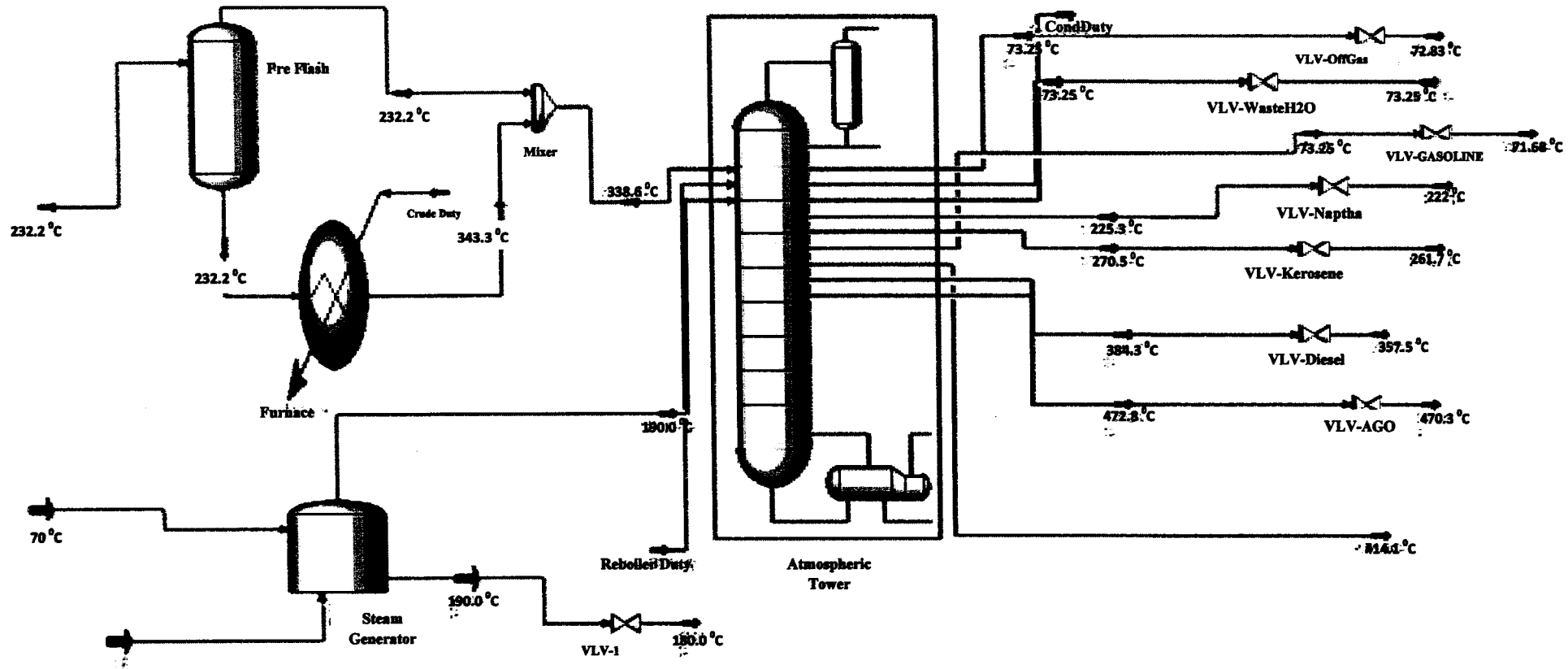


Figure 5.2: Temperature Flow Diagram

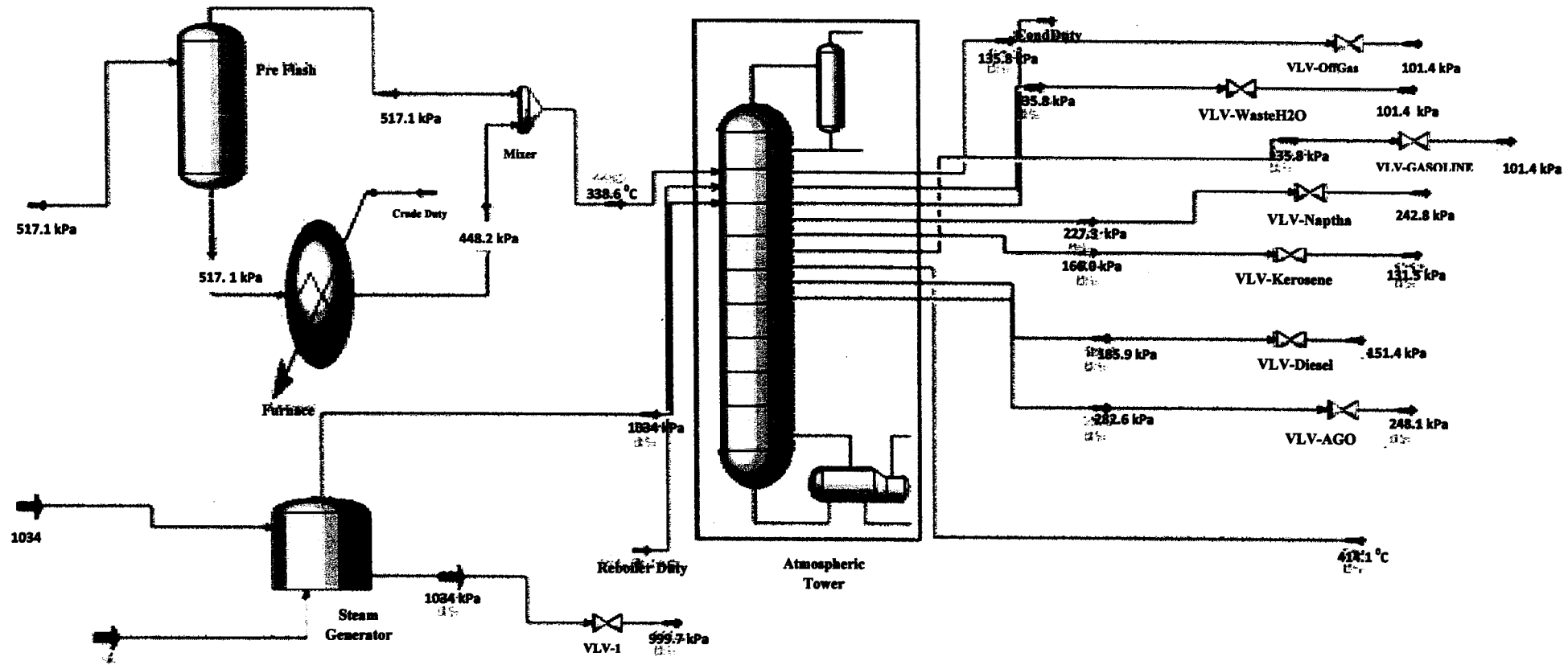


Figure 5.3: Pressure Flow Diagram

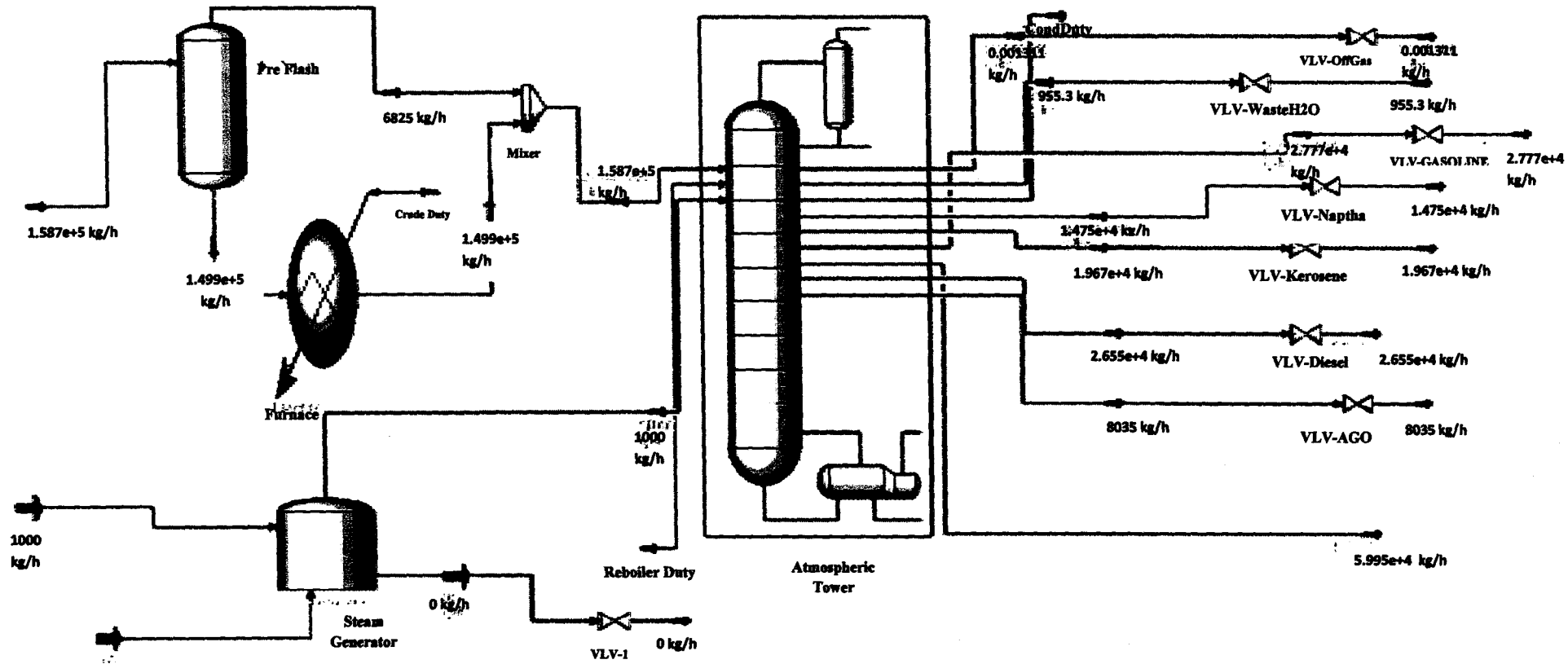


Figure: 5.4: Energy Flow Diagram

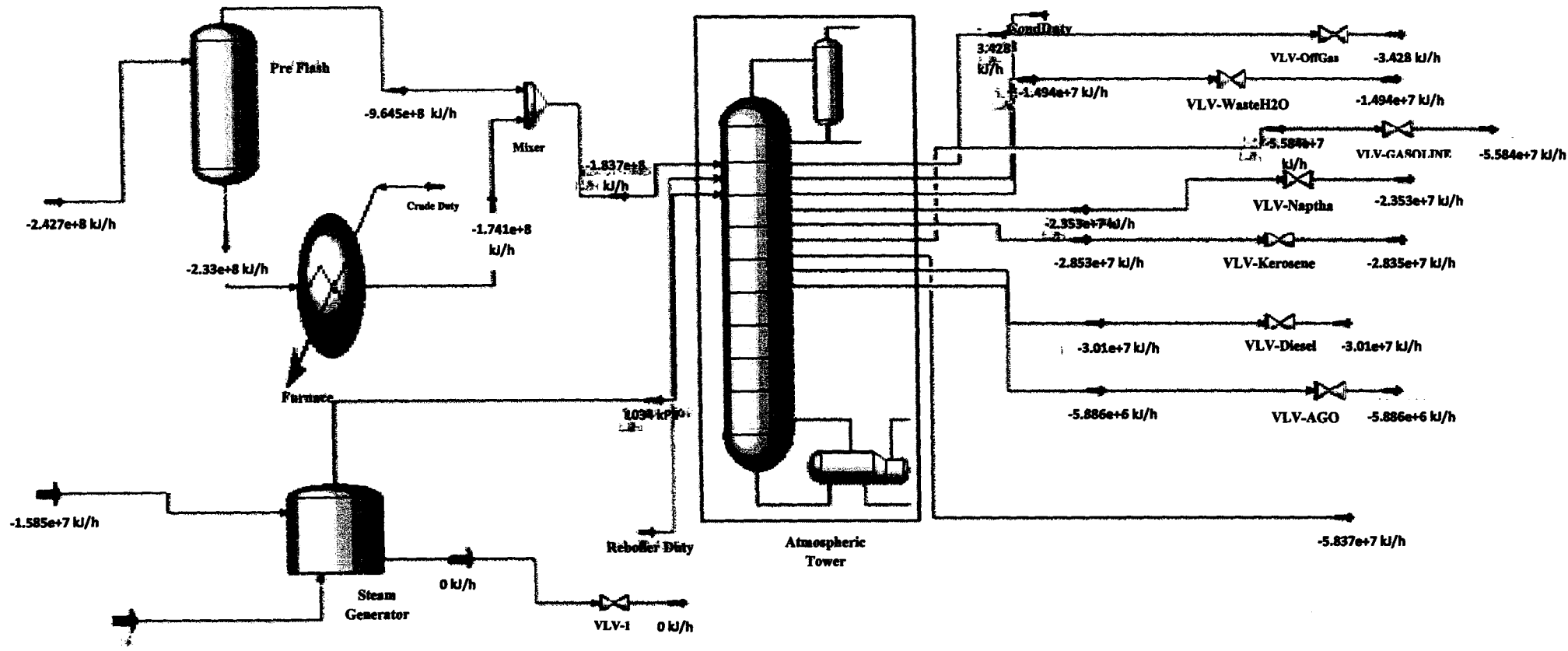


Figure 5.5: Mass Flow Diagram

Chapter Six

6.0 EQUIPMENT DESIGN

6.1 Separator (Preflash)

The separator (Preflash) used in this operation is a vertical one which is used in separating the vapour from the liquid phase before sending the crude oil to the furnace.

Generally, $V = vt$, where V is the vessel volume, v is the volumetric flow rate and t is the residence time. From chapter three, $v = 200 \text{ m}^3/\text{hr}$ and assuming a residence time of about 10.08 minute (0.168 hr), we have that:

$$\text{Vessel volume, } V = 200 \text{ m}^3/\text{hr} \times 0.168 \text{ hr} = 33.60 \text{ m}^3$$

Taking a Height to Diameter ratio of 5.5 (i.e. τ),

$$\Rightarrow \text{Height of vessel, } h = \tau d$$

$$\text{But } V = \pi d^2 h / 4$$

Where d and h are the diameter and height of the vessel respectively

$$\Rightarrow V = \pi d^2 \tau d / 4 = \pi d^3 \tau / 4$$

$$33.60 = (22/7 \times d^3 \times 5.5) \div 4$$

$$d = 1.981 \text{ m}$$

$$\therefore h = 5.5 \times 1.981 = 10.90 \text{ m}$$

6.2 Distillation Tower

The design of a distillation tower has always been a complex one involving a lot of trials and errors. The design of this tower therefore; since manual and not computer aided design calculations are done; has been limited to specifications such as diameter, height, efficiency and plate number.

From chapters three and four, we have the following parameters for the feed streams

Mass flow rate of Vapour, $M_v = 6825 \text{ kg/hr}$

Mass flow rate of Liquid, $M_l = 1.499 \times 10^5 \text{ kg/hr}$

Vapour Density $\rho_v = 687.8 \text{ kg/m}^3$

Liquid Density $\rho_L = 788.4 \text{ kg/m}^3$

Vapour Viscosity, $\mu_v = 1.119 \times 10^{-2} \text{ cP}$

Liquid Viscosity, $\mu_l = 0.3710 \text{ cP}$

Pressure, $P = 517.1 \text{ kPa}$

Volumetric flow rate, $V = \text{mass flow/density}$

$V_v = 9.9 \text{ m}^3/\text{hr}$; $V_L = 190.1 \text{ m}^3/\text{hr}$

Flow parameter, X is calculated as follows:

$$X = \frac{1.499 \times 10^5}{6825} \times \sqrt{\frac{687.8}{788.4}}$$

$$X = 20.51$$

Specifying the ratio of downcomer area to total area, $AdAt$

$$AdAt = 0.1 \text{ if } X \leq 0.1$$

$$0.2 \text{ if } X \geq 1$$

$$0.1 + \frac{X-0.1}{9} \text{ Otherwise } AdAt = 0.1$$

If X is smaller than 0.1, use $X=0.1$ in equation (4.31)

$$X = X \text{ if } X \geq 0.1$$

$$0.1 \text{ otherwise}$$

$$X = 0.1$$

Calculating the ratio hole to active area, A_h/A_a

$$AhAa = 0.907 \times \left(\frac{do}{p}\right)^2 \quad AhAa = 0.907 \times \left(\frac{4.5 \times 10^{-3}}{0.012}\right)^2$$

$$AhAa = 0.128$$

Given

$$D = \sqrt{\frac{4 \times QG}{f \times vGF(t) \times (1 - AdAt) \times \pi}}$$

Use is made here of the step function F(X) to define the recommended values of tray spacing.

$$t = 0.5 \times m \times (\phi(1 \times m - D) + 0.6 \times m \times (\phi(3 \times m - D) - \phi(1 \times m - D)) \dots$$

$$+ 0.75 \times m \times (\phi(4 \times m - D) - \phi(3 \times m - D)) \dots$$

$$+ 0.9 \times m \times (\phi(20 \times m - D) - \phi(4 \times m - D)) \dots$$

$$\left(\frac{D}{t}\right) = \text{Find}(D, t)$$

6.2.1 Diameter of column

$$D = 2.816 \text{ m} \quad t = 0.6 \text{ m}$$

Calculating some further details of tray design

$$At = \frac{\pi \times D^2}{4} \quad At = \frac{\pi \times 2.816^2}{4}$$

Total area

$$At = 6.229 \text{ m}^2$$

6.2.2 Atmospheric Distillation Column Number of Stages and Position of the Feed Stage Determination

In the distillation process at low temperatures differences between the top and the bottom of the column, we may suppose that the vaporization pressure of species may be constant in the whole apparatus. Because of that, we say

$$P_{vl} = \exp\left(18.5242 - \frac{3578.91}{303 + -50.50}\right) \times kPa \quad P_{vl} = 0.901 \text{ mmHg}$$

Vapor pressure of the heavier component (P_{vl} must be greater than p_{vh}).

$$P_{vh} = \exp\left(18.3036 - \frac{3818.44}{303 + -46.13}\right) \times kPa \quad P_{vh} = 0.365 \text{ mmHg}$$

And in the same way, let's define the pressure inside the column:

$$P = 1.01 \text{ bar}$$

With the statement above, we call the quotient as the relative volatility coefficient, which is evaluated as

$$\alpha := \left(\frac{p_{vl}}{p_{vh}}\right) \quad \alpha = \left(\frac{0.901}{0.365}\right)$$

$$\alpha = 2.47$$

With all the previous conditions, we may trace an equilibrium curve in a diagram y vs. x , where y is the composition of the lighter component in the gas phase and x is the lighter component composition in the liquid phase. This curve will have the next equation:

$$y_e(x_e) := \frac{\alpha x_e}{1 + (\alpha - 1)x_e} \dots \text{both } y_e \text{ and } x_e \text{ design "y is equilibrium with x"}$$

If the enthalpy of vaporization for both the heavier and lighter components is similar, the conditions of the McCabe-Thiele distillation are valid and we may use a "graphical" solution to the problem:

$$F = 1.577 \times 10^5 \text{ kg/hr}$$

$$z_f = 0.04:$$

6.1.4.1 Design Parameters

$X_d = 99.8 \text{ x } \%$...lighter component composition at the top

$X_b = 1 \text{ x } \%$...bottom composition of lighter component

$r = 0.99 \text{ x } \%$...lighter component recuperation in the process

$q = 0.98;$...thermodynamic constant of feed;

$$q: = \frac{Hv-Hf}{Hv-Hl}$$

Where Hv ...saturated vapor enthalpy at feed temperature.

Hl ...saturated liquid enthalpy at feed temperature.

Hf ...feed enthalpy.

$$\infty \cong 10^{100}$$

$$q: = q \times \left(1 + \frac{1}{\infty}\right)$$

6.1.4.4 Column efficiency:

$$\text{Col_Eff:} = 85 \%$$

$$\text{Actual_Number_of_Trays:} = \frac{Nt}{\text{Col_Eff}} \quad \text{Col_Eff} = 0.85$$

$$\text{Actual_Number_of_Trays} = \frac{16}{0.85}$$

$$\text{Actual_Number_of_Tray} = 19$$

6.1.4.5 Column Height

Column height = [Number of actual trays] x [Tray spacing] + Additional elevation of the liquid holdup in the base of the column.

$$\text{Tray_spacing:} = 0.55\text{m}$$

Additional elevation of the liquid holdup in the base of the column (Z_A)

$$Z_A: = 0.5\text{m}$$

Column height (H) is finally calculated as

$$\text{Column_Height:} = [(\text{Actual_Numbe_Trays}) \times \text{Tray_Spacing}] + Z_A$$

$$\text{Column_Height} = [(19) \times 0.5] + 0.5$$

$$\text{Column_Height} = 10\text{m.}$$

Chapter Seven

7.0 EQUIPMENT OPTIMIZATION

In optimizing any equipment, the first step is clearly to define the objective. That is, the criterion to be used to judge the performance of the system. In engineering design, the objective of optimizing any reactor or equipment must be an economical one. This is because for any chemical plant set up, the primary objective is to maximize profits.

7.1 Equipment Optimization

The optimization of the separator, which is used in separating the vapour from the liquid before sending it to the furnace, is carried out as shown below.

7.1.1 Optimization of the Separator

The Preflash can also be optimized using the fact that, in order to minimize cost of construction of the separator, the length and diameter of the tank must be kept at minimum.

Since the separator is a tubular type, it is said to have a cylindrical shape.

That is, the total surface area of the separator is given as

$$A_r = 2\pi r^2 + 2\pi rH$$

$$\text{and } r = \frac{D}{2}$$

where r = radius of the reactor tube

D = diameter of the reactor

H = height of the reactor

π = pie, a constant

So, the formula becomes

$$A_r = 2\pi r^2 + 2\pi rH$$

$$A_r = 2 \cdot \pi \cdot \left(\frac{D}{2}\right)^2 + 2\pi \cdot \frac{D}{2} \cdot H$$

$$A_r = 2 \cdot \pi \cdot \frac{D^2}{4} + 2 \cdot \pi \cdot \frac{D}{2} \cdot H$$

$$A_r = \pi \cdot \frac{D^2}{2} + \pi \cdot D \cdot H$$

From the above equation, one can now say that the area of the reactor is a function of the diameter and length of the reactor. That is, mathematically,

$$A_r = f(D,H)$$

Where $A_r = \pi \cdot \frac{D^2}{2} + \pi \cdot D \cdot H$ is the objective function and $D = \text{minimum}$ and $H = \text{minimum}$ are the constraints that are to be minimized, so that the equation of A_r becomes

$$f(D,H) = \pi \cdot \frac{D^2}{2} + \pi \cdot D \cdot H$$

Noting that the volume of the separator is given as

$$V_r = \pi \cdot r^2 \cdot H$$

$$\text{With } r = \frac{D}{2}$$

$$V_r = \pi \cdot \left(\frac{D}{2}\right)^2 \cdot H$$

$$V_r = \pi \cdot \frac{D^2}{4} \cdot H$$

Making H the subject of the formula in the above equation,

$$H = \frac{4 \cdot V_r}{\pi D^2}$$

Substituting the expression of H into the equation of total area of the separator which is a function of distance, D, and height, H, it is obtained that

$$f(D,H) = \pi \cdot \frac{D^2}{2} + \pi \cdot D \cdot \frac{4 \cdot V_r}{\pi D^2}$$

Simplifying,

$$f(D,H) = \pi \cdot \frac{D^2}{2} + \pi \cdot \frac{4V_r}{\pi \cdot D}$$

Now, it can be observed that the term of H has disappeared. That is to say that the total surface area is now a function of only the diameter, D. As such, the expression can be rewritten as

$$f(D) = \pi \cdot \frac{D^2}{2} + \pi \cdot \frac{4V_r}{\pi \cdot D}$$

Since the aim here is to optimize, it may either be maximizing or minimizing. In this case, the aim is to maximize but to minimize the dimension of the tank so that the size can be less, consequently, the profit can be much.

Differentiating the above equation,

$$\frac{d}{dD}f(D) = \pi \cdot D - \frac{4}{D^2} \cdot V_r$$

To optimize, the differential will be equal to zero, that is

$$\frac{d}{dD}f(D) = \pi \cdot D - \frac{4}{D^2} \cdot V_r = 0$$

Taking the last two expressions,

$$\pi \cdot D - \frac{4}{D^2} \cdot V_r = 0$$

Simplifying and making D the subject of the formula

$$\pi \cdot D = \frac{4}{D^2} \cdot V_r$$

$$D^2 \cdot D = \frac{4 \cdot V_r}{\pi}$$

$$D^3 = \frac{4 \cdot V_r}{\pi}$$

$$D = \sqrt[3]{\frac{4 \cdot V_r}{\pi}}$$

The above expression is now the optimized diameter of the reactor.

Using the relationship between the height of the reactor and the diameter given as

$$H = \frac{4 \cdot V_r}{\pi \cdot D^2}$$

Substituting for D in this expression yields

$$H = \frac{4 \cdot V_r}{\pi \cdot \left(\sqrt[3]{\frac{4V_r}{\pi}} \right)^2}$$

Simplifying,

$$H = \frac{V_r^{(1/3)}}{\pi^{(1/3)}} \cdot \sqrt[3]{2^2}$$

Numerically, with

$$V_r = 33.60 \text{ m}^3$$

$$D = \sqrt[3]{\frac{4 \cdot V_r}{\pi}}$$

$$D = 3.50 \text{ m}$$

$$H = \frac{V_r^{(1/3)}}{\pi^{(1/3)}} \cdot \sqrt[3]{2^2}$$

$$H = 4.41 \text{ m}$$

Thus, it therefore means that the optimum diameter and height of the separator are 3.50 m and 4.41 m respectively.

Chapter Eight

8.0 SAFETY AND QUALITY CONTROL

8.1 Safety

The legal measures present make it imperative for safety and good industrial practices to be enforced in order to safeguard the health of the employees and the general public, ensure efficient operation, increase productivity and prolong life of the plant. The major sources of hazards are with equipment handling of material, fire outbreaks, excessive light and heat, poor ventilation and plant layout in general. Some safety practices involved in the design and subsequent operation of this plant include;

1. The injector head should be covered with a labelled casing to prevent contact with workers.
2. Provision of automated control systems, interlocks, alarms and trips.
3. Fire hydrants, extinguishers, smoke alarms and fire exits will be provided at vantage points.
4. Overalls and gloves, nose masks and proper waste disposal.
5. Earplugs will be provided for all plant personnel.
6. All equipments should be clearly coded and labelled to inform plant personnel on fluids flowing through them.

The processing area is the most dangerous area of the plant because this area is full of mechanical equipments which are subject to failure. Processing units should be consolidated rather than scattered in order to ease their identification as dangerous areas. Concrete floors will be generally used, with the floor made to slope gently where necessary to prevent pools of water on the floor. All the factory workers should be thought or educated about safety rules and precautions. They should be introduced to various industrial hazards in every section of the factory. Newly employed workers should be introduced to the details of the factory production process and be guided on the safety regulations. They should be taught on how to render first aid attention whenever an accident occurs, they should also be enlightened on the use of individual protective facilities.

A worker should only be permitted to work, if he had been sufficiently inducted on the major safety regulations and unit operations in the factory, he must be acquainted with the new safety regulations and procedures (Odigure, 1998)

Environmental and safety concerns mean that oil refineries are sometimes located some distance away from major urban areas. Nevertheless, there are many instances where refinery operations are close to populated areas and pose health risks such as in the Campo de Gibraltar, a CEPSA refinery near the towns of Gibraltar, Algeciras, La Linea, San Roque and Los Barrios with a combined population of over 300,000 residents within a 5-mile (8.0 km) radius and the CEPSA refinery in Santa Cruz on the island of Tenerife, Spain which is sited in a densely populated city center and next to the only two major evacuation routes in and out of the city. In California's Contra Costa County and Solano County, a shoreline necklace of refineries, built in the early 1900s before this area was populated, and associated chemical plants are adjacent to urban areas in Richmond, Martinez, Pacheco, Concord, Pittsburg, Vallejo and Benicia, with occasional accidental events that require "shelter in place" orders to the adjacent populations.

Because of their inherent hazards, especially from explosion, fire, and chemicals, oil refineries are tightly regulated places in which to work. Work permits must always be obtained and followed. Plant practices, warnings, and emergency procedures must be observed at all times. When in doubt, remember to exercise your "right to know" under WHMIS legislation. Check labels and MSDSs. Any required protective equipment and procedures must be explained and available when hazardous exposure is possible.

The plant and equipment of refineries are generally modern, and the processes are largely automatic and totally enclosed. Routine operations of the refining processes generally present a low risk of exposure when adequate maintenance is carried out and proper industry standards for design, construction, and operation have been followed. The potential for hazardous exposures always exists, however. Because of the wide variety of hydrocarbon hazards and their complexity, it is impossible to identify all of the hazards

here – and impossible for construction crews to know everything they may need for protection when performing maintenance, repair, or installation work in an oil refinery.

As a worker you must depend on the knowledge available from the plant operating and maintenance staff, normally available through your employer. If there is reasonable doubt about a situation in which you find yourself, exercise your “right to know” and make use of WHMIS to obtain the information, equipment, and procedures necessary to protect yourself and your fellow workers.

8.1.1 Hazardous Chemicals

In a refinery, hazardous chemicals can come from many sources and in many forms. In crude oil, there are not only the components sought for processing, but impurities such as sulphur, vanadium, and arsenic compounds. The oil is split into many component streams that are further altered and refined to produce the final product range. Most, if not all, of these component stream chemicals are inherently hazardous to humans, as are the other chemicals added during processing. Hazards include fire, explosion, toxicity, corrosiveness, and asphyxiation. Information on hazardous materials manufactured or stored in a refinery should be supplied by the client's representative when a work permit is issued.

8.1.2 Fire and Explosion

The principal hazards at refineries are fire and explosion. Refineries process a multitude of products with low flash points. Although systems and operating practices are designed to prevent such catastrophes, they can occur. Constant monitoring is therefore required. Safeguards include warning systems, emergency procedures, and permit systems for any kind of hot or other potentially dangerous work. These requirements must be understood and followed by all workers. The use of matches, lighters, cigarettes, and other smoking material is generally banned in the plant except in specially designated areas.

8.1.3 Major Shutdown and Maintenance

The principal exposures to hazardous substances occur during shutdown or maintenance work, since these are a deviation from routine operations. Plant turnarounds require careful planning, scheduling, and step-by-step procedures to make sure that unanticipated

exposures do not occur. Any plant shutdown requires a complete plan in writing to cover all activities.

8.1.4 Safe Work Practices and Procedures

8.1.4.1 Personnel

- Wearing protection and safety glasses must be worn in all operating areas or as posted.
- Respiratory protection or equipment must be fit tested.

Facial hair is unacceptable where the mask must make an airtight seal against the face.

- Shirts must be long-sleeved and worn with full-length pants or coveralls.
- Clothing must not be of a flammable type such as nylon, Dacron, acrylic, or blends. Fire-resistant types include cotton, Nomex, and Proban.
- Other PPE required may include acid hood, impervious outerwear, rubber boots, face shields, rubber gloves, disposable coveralls, monogoggles, and fall-arrest equipment.
- Smoking is allowed only in designated areas.

8.1.4.2 Vehicles

- Vehicle entry is by permit only and keys are to be left in parked vehicles.
- Vehicles must be shut down at the sound of any emergency alarm.
- Vehicles must be equipped with ground straps or cables.

8.1.4.3 Permit Systems

No work takes place in a refinery without a safe work permit. A safe work permit is a document issued by an authorized representative of the client permitting specific work for a specific time in a specific area. Work permits should indicate the date and time of issue, the time of expiry, a description of the work to be done, and the name of the company performing the work. Permits also specify any hazards and controlled products under WHMIS and any protective equipment needed for the job. The permit will advise you of any steps required to make the area or equipment safe for work, tell you the results of any gas tests, advise you of any electrical lockouts that have been done, and tell you of any work practices required for the specific job.

Safe work permits are valid only for a limited time and must be renewed following expiry or normally after any one-hour stoppage, after an emergency warning on the site, or for

other safety reasons. After such an event, any required gas testing or other testing must be repeated to ensure a safe return to the work. The types of safe work permits required typically include the following (specific categories may vary from site to site):

Hot work: covers any work that involves heat or an ignition source, including welding, grinding, and the use of any kind of motor. In high-risk areas, a spark watch may be required.

Benzene: required when a benzene exposure hazard exists.

Confined space entry hot work: involving potential ignition hazards.

Confined space entry cold work: involving work that will not produce a spark.

Electrical: for other than routine work.

Camera: typically requires a hot work permit when lighting is required.

Asbestos: required whenever an asbestos exposure hazard exists.

11.Hydrant – permits the use of plant fire hydrants.

8.1.4.4 Special Authorization Permits

In addition to safe work permits, special authorization permits are normally required for the following operations:

- excavation
- hoisting with major mobile equipment
- hot tap and non-conventional repairs
- opening live flare lines
- temporary electrical facilities.

8.1.4.5 Emergency Warning System and Procedures

In oil refineries there will be both plant alarms or whistles and individual unit alarms. All workers must receive training in recognizing and responding to these alarms. Verbal messages usually accompany the alarms. There will be different alarms for a fire emergency and toxic alarms.

- When an alarm sounds, secure all equipment and shut down all vehicles.
- Note the wind direction (wind socks) and proceed to the appropriate assembly area (or safe haven).

- Do a head count to make sure all personnel are accounted for and report the result to a client contact person.

- Know the local designated safety areas or safe havens and emergency phone number(s).

If you are the one who is first aware of an emergency, then call the emergency number.

- Report your name.

- Describe the emergency.

- Identify its location.

- Indicate whether anyone is injured.

- Proceed to the assembly area.

8.1.4.6 *Electrical Precautions*

- Electrical tagging and lockout procedures must be understood and followed by all workers.

- All electric tools, cords, and equipment must be grounded or double-insulated.

- Use explosion-proof fixtures where required.

8.1.4.7 *Sewers*

- Sewers must be covered when hot work is being done in the vicinity.

- Sewer covers must be in good condition with no openings for vapor flow.

- Sewer covers are to be removed when hot work is discontinued at the end of the job or overnight to accommodate drainage.

8.1.4.8 *Blinding or Blanking-off*

- Piping connected to a work area from vessels, pumps, and other sources is isolated or blinded with a solid plate prior to the start of work.

- Blanking can sometimes be done with two valves and a bleeder valve between them. In this case the valves should be closed, chained, locked, and tagged.

8.1.5 Pollution Control

The refining process releases numerous different chemicals into the atmosphere; consequently, there are substantial air pollution emissions and a notable odor normally accompanies the presence of a refinery. Aside from air pollution impacts there are also

wastewater concerns, risks of industrial accidents such as fire and explosion, and noise health effects due to industrial noise.

The public has demanded that many governments place restrictions on contaminants that refineries release, and most refineries have installed the equipment needed to comply with the requirements of the pertinent environmental protection regulatory agencies. In the United States, there is strong pressure to prevent the development of new refineries, and no major refinery has been built in the country since Marathon's Garyville, Louisiana facility in 1976. However, many existing refineries have been expanded during that time. Environmental restrictions and pressure to prevent construction of new refineries may have also contributed to rising fuel prices in the United States. Additionally, many refineries (over 100 since the 1980s) have closed due to obsolescence and/or merger activity within the industry itself. This activity has been reported to Congress and in specialized studies not widely publicized.

A pollution type realized in the plant is that of water. Waste water from the entirety of the plant will be filtered to remove dirt and other solid particles after which the water is recycled back into the cooling tank or otherwise into drainage.

The unavoidable noise levels generated by some of the process units will cause noise pollution. Ear plugs will be worn by all workers operating any heavy equipment and workers around vicinity of loud noise to prevent the harmful effect of noise on the human ear. Sound absorbing materials will be employed in constructing all plant buildings that house noise generating equipments, thus reducing significantly the noise levels in these parts of the plant.

8.2 Quality Control

Quality control is a process by which entities review the quality of all factors involved in production. This approach places an emphasis on three aspects

1. Elements such as controls, job management, defined and well managed processes, performance and integrity criteria, and identification of records
2. Competence, such as knowledge, skills, experience, and qualifications
3. Soft elements, such as personnel integrity, confidence, organizational culture, motivation, team spirit, and quality relationships.

The quality of the outputs is at risk if any of these three aspects is deficient in any way. Quality control emphasizes testing of products to uncover defects, and reporting to management who make the decision to allow or deny the release, whereas quality assurance attempts to improve and stabilize production, and associated processes, to avoid, or at least minimize, issues that led to the defects in the first place.

8.2.1 Quality control measures

- (i) The need of the consumer: this is the most significant measure that must always be made to influence cost analysis and the production overall efficiency. The color, texture and the weight of the final product should be based on the demand of the customers who will determine the suitability of the product for further utilities.
- (ii) Adhesion to design specification as seen by the management.
- (iii) Selection of tolerance is done for quality characteristics.
- (iv) There is need for review of design documentation from time to time to ensure that the variation occurring the changing conditions is checked; the changes are released.
- (v) Supply performance history is established since the previous history is a record of demonstrating capability, rather than potential capability.
- (vi) For every equipment newly purchased, a check is carried out to ensure that the design is consistent with existing machines and process capabilities.

The inspections carried out include:

- (a). Specific guidelines of the characteristics to be checked.
- (b). Detailed description of special techniques or methods which are required.
- (c). Instructions for use.

Chapter Nine

9.0 PROCESS CONTROL AND INSTRUMENTATION

9.1 Introduction

In the design of an industrial plant, the methods which will be used, for plant operation and control help determine many of design variables, for examples, the extent of instrumentation can be factor in choosing the type of process and setting the labour requirements. It should be remembered that maintenance work would be necessary to keep the installed equipment and facilities in good operating condition. The importance of such factors which are directly related to plant operation and control must be recognized and take into proper account during the development of a design project.

Processes may be controlled more precisely to give more uniform and higher-quality products by the application of automatic control, often leading to higher profits. Therefore, any process with an input and output which may be flow, pressure, liquid level, temperature, composition or any inventory, environmental or quality variable that is to be held at a desired value must have some measure of control applied to it. Change in output may occur:

1. Random as caused by changes in weather or raw material quality
2. Diurnally with ambient temperature
3. Manually when operators change production rate
4. Stepwise when equipment is switched in or out of service, or
5. Cyclically as a result of oscillation in other control loops

Variation in any of the ways stated above would drive the output (controlled variable) further away from the set point (desired value) thus requiring a corresponding variable to bring it back (manipulative variable) .

There are various control methods used in chemical engineering such as feed forward control, feedback control system for a condenser (heat exchanger) in figure 9.1

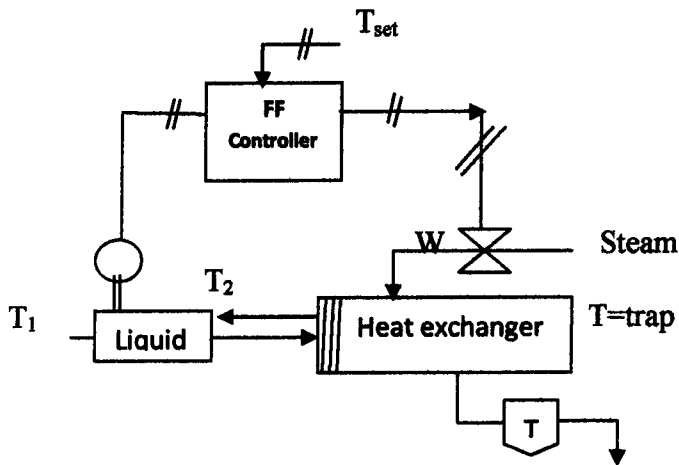


Figure 9.1: Feed forward control for a shell and tube heat exchanger

9.2 Column Control

9.2.1 Control of feed, high purity distillate composition, bottoms temperature

This control scheme is straightforward. This is done by first putting a flow controller on the feed. Since it is high purity overheads, the composition is regulated by adjusting the distillate rate. This is achieved with temperature in the top half of the column but some distance down as an inner loop cascaded with an analyser. Bottoms temperature is controlled with the steam rate. Finally the inventory loops - reflux drum level controlled by the reflux rate and the reboiler level on the bottoms rate.

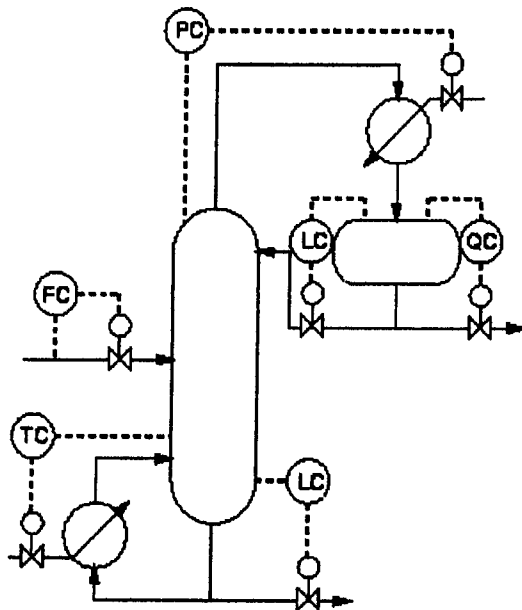


Figure 9.2: Control of feed, high purity distillate composition, bottoms temperature

9.2.2 Control of distillate rate, distillate composition, bottoms composition

Distillate rate is set by flow control, so, reflux must be used for drum level. Overheads composition can be regulated by boil up rate, which is quite satisfactory because of the low, vapour inventory in the column. But the bottoms composition could be done on bottoms flow, but reboiler level would then have to be by feed adjustment. Unfortunately this has a long series of hydraulic delays from tray to tray. It might be better to do reboiler level with bottoms and bottoms composition by feed, although this is an unusual scheme!

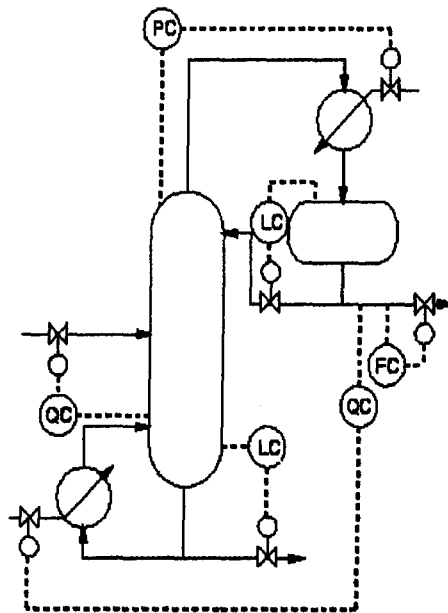


Figure 9.3: Control of distillate rate, distillate composition, bottoms composition

9.2.3 Control of distillate rate, bottoms rate, bottoms composition

Both product rates can be controlled by flow control. The level in the reflux drum can be controlled by the reflux rate. The bottoms composition is controlled by the reboiler steam rate. This leaves the level in the reboiler either floating or unsatisfactorily controlled by the feed rate.

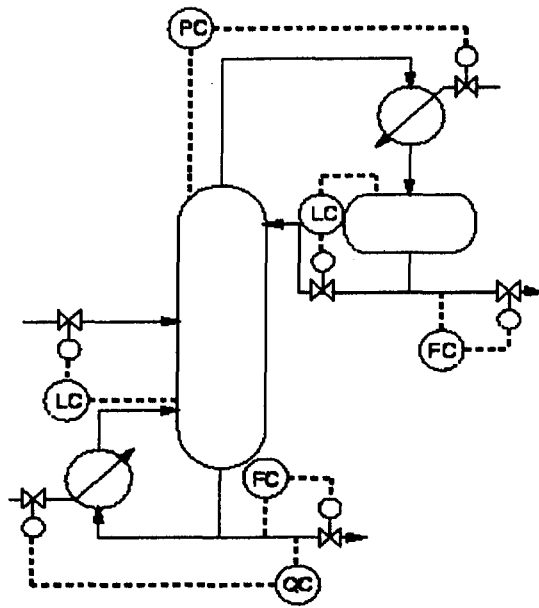


Figure 9.4: Control of distillate rate, bottoms rate, bottoms composition

9.2.4 Control of bottoms rate, distillate composition, bottoms composition

Distillate composition controlled conventionally either by distillate or reflux rate depending on purity, the other adjustment used for reflux drum level. Bottoms composition is by reboiler steam rate leaving the same difficulty with reboiler level as before.

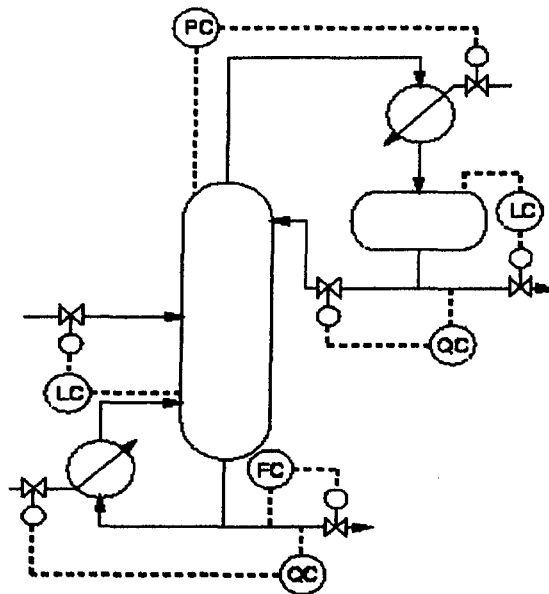


Figure 9.5: Control of bottoms rate, distillate composition, bottoms composition

9.3 Specification of Controller/Instrument Type

From the control diagrams shown above, the different controllers and instruments used are:

- TI Temperature Indicator [thermocouples (Chromel-alumel)]
- TC Temperature Controller (PID – Proportional-Integral-Derivative controller)
- PC Pressure Controller (P – Proportional controller)
- PI Pressure Indicator (Differential Pressure Gauge - diaphragm)
- LC Level Controller (PI – Proportional-Integral controller)
- LI Level Indicator (Float-type indicator)
- FC Flow Controller (PI – Proportional-Integral controller)
- FI Flow Indicator (orifice plate)

---- Pneumatic control due to high flammability of the process fluid

9.4 Instrumentation and Control and Control Objectives

The key objectives adhered to in the specification of the instrumentation and control schemes are:

1. Safe Plant Operations

- i. To keep the process variables within known safe operations limits
- ii. To detect dangerous situations as they develop and to provide alarms and automatic shut-down systems
- iii. To provide interlocks and alarms to prevent dangerous operation procedures.

2. Production Rate: To achieve the design producer output

3. Product Quality: To maintain the product composition within the specific gravity standards

4. Cost: They operate at the lowest production cost, but not to the detriment of the product quality

In the plant design some of the variables needed to be monitored and controlled are, the flow rate, temperature, pressure and composition.

- i. **Flow Rates/Feed Ratio control:** a feed ratio controller is applied between the crushed preheat crude, Preflash and preflash liquid with the aim of maintaining a complete refining crude oil to the desired products
- ii. **Temperature Control Reaction:** temperature is controlled by regulating the flow rate reaction into it. This can also be controlled by heat transfer. The reaction is exothermic. It produces heat, which tend to raise reaction temperature thereby increasing reaction rate and producing more heat. This positive feedback is countered by negative feedback in cooling system, which removes more heat as reactor temperature rises. The temperature controller, which in turn operates the coolant valve, to counter the rise or drop in temperature in the sector.
- iii. **Pressure Control:** pressure sensing is quite straight forward with the aid of pneumatic instrumentation such as Bourdon gauge, diaphragms and bellows. These sensors measure absolute pressure and pressure differences between two levels. Therefore, pressure control is achieved by manipulating the airflow rate in the compressor to avoid deviation from set point.
- iv. **Composition Control:** First requirement here is to establish proper stoichiometry of the reactants in proportion needed to satisfy the reaction chemistry and also desired output product. This is achieved by setting input flow rates in ratio to one another, or a composition measurement (analyzer) can be used to trim the ratios to the right proportion.

9.5 Operability Study

An “operability study” is a structured technique for identifying potential mal – function beforehand. Critical examination in process design, inevitably leads to a safer, more reliable and more profitable process.

These studies are based on sets of “keywords” which stimulates thoughts. It is essential to formulate unambiguously the design objectives, to which the key words are applied in a systematic way.

In operability studies, the thoughts are about deviations from the design conditions and the design intention. The set of key words contains two sub-sets: “property words” which focus attention on the design conditions and the design intention; and the second set of “guide words” which focus attention onto possible deviations. The Chemical industries Association has published a Guide to Hazard and Operability Studies which recommends the seven guide words explained. (Douglas, 2004)

Table 9.1: A List of Guide Lines

NO or NOT	The complete negation of these intentions	No part of the intentions is achieved but nothing else happens.
MORE, LESS	Quantitative increases or decreases	These refer to quantities and properties such as flow rates and temperatures as well as activities like “ABSORB” and “REACT”.
AS WELL AS	A qualitative increase	All the design and operating intentions are achieved together with some additional activity.
PART OF	A qualitative decrease	Only some of the intentions are achieved, some are not.
REVERSE	The logical opposite of the intention	This is mostly applicable to activities, for example, reverse flow or chemical reaction.
OTHER THAN	Complete substitution	No part of the original objective achieved; different thing happen.

Efficient to use only NO, MORE and LESS, with a layer set of property words should be chosen beforehand accordingly. (Douglas, 2004)

Operability Studies are generally carried out in the industry by a small team of specialists, directed by an experienced group- leader. The exercise is of the brain – storming type and only those deviations which would lead to hazardous outcomes are recorded for further action. Initially the experienced chemical engineer will find it more effective to record all causes of deviations, their consequences and action. In addition, it is more appropriate to

use a process follow-sheet with only the basic control loops. This enables those who are not experienced instruments which are included in the final diagram. A further benefit of keeping complete records of the operability study is that operating instructions and procedures for start-up, shut-down and carrying out cyclic operations can be drawn up by reference to the operability study. Furthermore, if significance of deviations in measured variables is understood, simple fault-finding strategies can be included in the design report. The operability study in this design project differs from that of an industrial process design in that, normally every vessel and pipeline must be covered by an industrial operability study. In this design report, the operability study is restricted to the scrubber section only. (Douglas, 2004)

Chapter Ten

10.0 ENVIRONMENTAL ACCEPTABILITY

10.1 Identification of Possible Pollutants

1. Air Emissions.
2. Liquid Effluents (Wastewater).
3. Solid and Hazardous Waste
4. Ambient Noise

Refining process releases numerous different chemicals into the atmosphere; consequently, there are substantial air pollution emissions and a notable odor normally accompanies the presence of a refinery. Aside from air pollution impacts there are also wastewater concerns, risks of industrial accidents such as fire and explosion, and noise health. Waste Characteristics from Boilers, process heaters, and other process equipment are responsible for the emission of particulates, carbon monoxide, nitrogen oxides (NO_x), sulfur oxides (SO_x), and carbon dioxide.

Sulfur recovery units, combustion units and flares release SO_x . Catalytic cracking regenerators release particulates, NO_x , and SO_x . Catalyst change over's and cokers release particulates. Volatile organic compounds (VOCs) such as benzene, toluene, and xylene are released from storage, product loading and handling facilities, oil/water separation systems, and as fugitive emissions from flanges, valves, seals, and drains. For each ton of crude processed, refineries may emit about:

0.8 kg (ranging from less than 0.1 to 3 kg) of particulate matter;

1.3 kg of SO_x (ranging 0.2-6 kg and 0.1 kg with Claus sulfur recovery process);

0.3 kg of NO_x (ranging 0.06-0.5 kg);

2.5g of BTX (benzene, toluene, xylene) (ranging from 0.75 to 6) and 1g with Claus sulfur recovery process. Of this, about 0.14g of benzene, 0.55 g of toluene, and 1.8 g of xylene may be released per ton of crude processed. VOC emissions depend upon the production

techniques, emission control techniques, equipment maintenance, and climate conditions and may be 1 kg per ton (with a range of 0.5 to 6 kg/t) of crude processed.

Petroleum refineries use relatively large volumes of water especially for cooling systems. Surface water run-off and sanitary wastewaters are also generated. The quantity of wastewaters generated and its characteristics depend on the process configuration. As a general guide, approximately 3.5-5 m³ of wastewater per ton of crude is generated when cooling water are recycled. Refineries generate polluted wastewaters, containing BOD₅ and COD levels of approximately 150-250 mg/L and 300-600 mg/L, phenol levels 20-200 mg/L; oil levels of 100 to 300 mg/L in desalter water and up to 5,000 mg/L in tank bottoms; benzene levels of 1 to 100 mg/L, benzo(a)pyrene level of less than 1 to 100 mg/L, heavy metals (chrome and lead levels of 0.1-100 and 0.2-10 mg/L respectively), and other pollutants. The refineries also generate solid wastes and sludges (with a range of 3-5 kg per ton of crude processed), 80% h effects due to industrial noise.

10.1.1 Air Emission

Air emission sources at petroleum refineries generally include:

- Boilers, Engines, and Turbines. Boilers are used to produce steam and to raise the temperature of feed materials to meet reaction or distillation requirements. Boilers are typically fired with residual oil, distillate oil, refinery gas, or natural gas. CO-rich regenerator flue gas also may be used as fuel. Reciprocating engines are used to drive gas compressors and are usually fueled by natural gas or refinery gas. Gas turbines can fire a variety of fuels and are used as cogeneration units that produce electricity and steam for process needs.

10.1.1.1 Process Heaters

Process heaters are used to indirectly heat process fluids to a moderate processing temperature. Fuel is combusted in an insulated chamber and the heat transferred to either tubes or a cylindrical vessel containing the process fluid. Process heaters use natural gas, refinery fuel gas, fuel oil or residual oil as fuel.

10.1.1.2 *Blowdown Systems*

Blowdown systems permit the removal of liquids and vapors from process units in order to permit shutdown of the process unit for maintenance/repair or to prevent dangerous high temperature or high pressure conditions from occurring. A refinery blow down system consists of valves, piping, surge vessels, etc. to allow safe transfer of process liquids or vapors out of each process unit. Because this transfer can allow the release of very concentrated air emissions, the blowdown system is usually connected to a refinery flare unit in order to destroy these high concentration air emissions.

10.1.1.3 *Process Equipment Leak*

Emissions occur from process equipment whenever components in the liquid or gas stream leak. Components such as pumps, valves, pressure relief valves, and flanges are potential sources that can leak due to seal failure.

10.1.1.4 *Wastewater Treatment*

All refineries employ some form of wastewater treatment prior to discharge to the environment or reuse in the refinery. Many of these collection and treatment system units are open to the atmosphere and allow organic-containing water to contact ambient air. Emission sources from wastewater collection and treatment systems include process drains and collectors, oil-water separators, air flotation systems, and surface impound basins and ponds.

10.1.1.5 *Cooling Water Towers*

Refineries use large quantities of water for cooling throughout the refining process. Cooling towers are used to transfer heat from the cooling water to the atmosphere. Water that enters the tower may contain hydrocarbons from leaking equipment.

10.1.2 *Liquid Effluents*

Refinery wastewaters often require a combination of treatment methods to remove oil and other contaminants before discharge. Separation of different streams (such as storm water, cooling water, process water, sanitary, sewage, etc.) is essential to minimize treatment requirements. A typical system may include sour water stripper, gravity oil/water separation, dissolved air flotation, biological treatment and clarification. A final polishing

step using filtration, activated carbon, or chemical treatment may also be required. Achievable pollutant loads include: BOD₅ of 6 g, COD of 50 g, suspended solids of 10 g and oil and grease of 2 g, all per ton of crude processed.

10.1.3 Solid and Hazardous Wastes

Sludge treatment is usually performed using land application (bioremediation), or solvent extraction followed by combustion of the residue or used in asphalt, where feasible. In some cases, the residue may require stabilization prior to disposal to reduce the leachability of toxic metals. Oil is recovered from slops using separation techniques such as gravity separators and centrifuges.

10.2 Treatment of Possible Pollutants

1. Air Emissions can be treated by using the following:
 - Minimize losses in storage tanks and product transfer areas by methods such as vapor recovery systems and double seals.
 - Minimize sulphur oxides emissions either through sulphur desulphurization of fuels to the extent feasible or by directing the use of high sulphur fuel to units equipped with sulphur oxides emissions control.
 - Use low nitrogen oxide burners to reduce nitrogen oxide emissions.
 - Keep fuel usage to a minimum.
2. Liquid Effluents such as waste water can be treated using the following methods:
 - Air is mixed vigorously into tanks with waste water effluent.
 - Chlorine or bleach is added to the waste water effluent.
 - Sand and grit are removed from the waste water effluent.
 - Bacteria are settled out of waste water effluent.
 - Oils and greases float to surface of waste water effluent.
3. Solid and Hazardous Waste: sludge treatment is usually performed using land application (bioremediation) or solvent extraction followed by combustion of the residue or by use for asphalt, where feasible. In some cases, the residue may require stabilization prior to disposal to reduce the leach ability of toxic metals.

Chapter Eleven

11.0 START UP AND SHUTDOWN PROCEDURE

Start up time may be defined as the time span between end of construction and the beginning of normal operations (Odigure, 1998). Start up and shutdown procedure must proceed safely and be flexible enough to be carried out in various ways. In other words, the start up and shutdown of the atmospheric distillation unit should be such that it can be easily and safely operated. The operating limits of the plant should not be exceeded and dangerous mixtures must not be formed as a result of abnormal states of concentration, composition, temperature, phase, pressure, reactants and products.

It should be noted that some items of actions must be completed before even the start up of the plant in order to prepare the plant for the start up operation.

1. All scaffolds and temporary piping and supports should be removed.
2. Lines and equipment should be flushed out.
3. Pumps, motors/turbines and compressors should be run.
4. Hydrostatic or pneumatic lines and equipment should be tested.
5. Laboratory and sampling schedule should be prepared.
6. All instruments should be inspected and tested (Odigure, 1998).

11.1 Start up Procedures

For the Atmospheric distillation unit designed, the following are the start up procedures:

1. The boiler should be started up using auxiliary fuel to begin steam production.
2. The reactor should be heated up and maintained at their operating temperatures and pressures.
3. The inlet and outlet valves of the reactors should be opened up.
4. The compressor valve to supply air at a regulated pressure should be opened.
5. All the inlet and outlet valves to the reactor should be shut as soon as the feed enters the reactor (ADU) in order to achieve maximum conversion.
6. The outlet valve of the reactor should be opened for the evacuation of the reactor contents.

7. The outlet valve of the reactor should be locked before opening the inlet valve before further entering of the feed into the reactor.

11.2 Shutdown Procedures

The shutdown procedures for the Atmospheric distillation unit are as outlined below:

1. The supply of the pre-heated crude to the preflash column should be cut off.
2. The water supply into the steam generator should also be cut off.
3. All the purge valves should be opened to discharge unconverted reactants.

11.2.1 Emergency Shutdown of Plant

Emergency shutdown of the configuration is deemed necessary if there is abnormality, which might result in costly top (unwanted) event. This may be due to breakdown of any major equipment that is not easily replaceable. For example, if the steam generator has a major problem or breakdown completely, there is no other option than to shut down the atmospheric distillation unit. Insufficient utility services may also necessitate emergency shutdown of plant.

When there is an emergency the process trip system is designed to affect the emergency shutdown of the plant. Unit shutdown is preferable; it would not affect production and quality because of the cost involved in starting up the whole unit.

A trip system carries out the appropriate activity on command from automatic actuation of the relay or other means. Such systems are closely associated to the shutdown of plant when some units are operating on abnormal manner. The shut down can be designed to subdivide the plant into different segments. These segments include: operating utilities, pressure level, or volume of flammable material. Thus, it would be possible to reduce the volume of fluid that would be spilled in the event of pipe rupture.

This system must be reliable and respond only when required, so as to avoid unnecessary shut down of plant and also when necessary they must not fail. The trip setting should be designed to incorporate safety.

11.2.2 Start Up after Emergency Shutdown

When the trip system is being designed, each start up must be undertaken with care. It is necessary to consider what happens when the trip system occurs and when the trip system condition is removed. The necessary actions are planned for all those cases, which cannot be foreseen.

In addition, during the start up after emergency shutdown,

1. Undesirable or toxic materials under pressure should be removed during the start up.
2. Non-conforming specification process materials should be recycled.
3. Air and undesirable materials or chemicals used for cleaning the system should be removed.
4. The controls and set points should be checked and adjusted.
5. It should be ensured that everything is in good order before the start up.

Chapter Twelve

12.0 SITE FOR PLANT LOCATION

Selecting a plant site is critical to the financial success of a plant. Several factors must be considered in selecting a general plant site location. The procedure for choosing a specific plant location can be presented in a series of required steps. After the site is selected, factors that go into getting the facility built, including permitting and the other necessary legal steps, need to be considered.

12.1 Site Factors

The primary siting factors that influence the selection of a plant location are as follows: environmental and safety considerations, labor availability and productivity, raw material availability, proximity to market, property cost, accessibility to transportation, tax incentives, electric power availability and cost, and living conditions.

In the selection of a plant site, it is a good idea to get broad-based input, including information from sales, production, plant engineering, and from the general manager. The first objective is to narrow the range of possible choices. This involves focusing on the most important criteria, which differ widely for each type of facility.

One of the key decisions that can influence site selection is whether to use conventional or modular construction. Such a decision will have an enormous economic impact on the project, since vastly different technical considerations and types of investments will be required for each approach. Conventional or “stick built” construction strategies are time honoured: the engineering, procurement, and construction are done in a logical, consecutive fashion, with some parts done in parallel with the goal being to execute the project as expediently as possible. Equipment and materials are procured from worldwide suppliers and are delivered to the site. At the site, roads are paved, equipment is set, piping is erected and electrical wiring is completed, all according to the drawings, specifications and standards developed during the detailed engineering phase of the project.

However, at a particular site, the conventional construction approach may not be the most cost effective method for project execution. Factors that directly or indirectly affect

construction project execution—and may favor one design and construction approach or site selection over another—may include construction labor availability at the site, weather conditions, the logistics of transporting equipment and materials to the site, schedule considerations and permitting issues.

Modular design and offsite construction of the plant may help to overcome many obstacles, such as weather and scheduling conflicts. In general, modularization involves the assembly of a plant at a convenient offsite location, using either local or imported labor. Each module represents a completed section of a plant including equipment, piping, electrical, instrumentation, insulation and painting. There has to be a methodology for transporting the modules to the site. Modularization may render a site economical and practical whereas it may not otherwise be with conventional construction.

12.2 Environmental and Safety Considerations

No matter how advantageous a site location may be, if a permit to build cannot be obtained or the uncertainties in getting the necessary permits jeopardize the timing of a project, then it may be necessary to choose another site. Thus, environmental considerations may be, overall, the most important siting factor.

12.3 Labor Availability and Productivity

Plants need to be run by people and the availability of employees can constitute the overriding consideration in certain businesses, in relation to siting. Labor-intensive businesses have to either move to a location where labor is available or move their employees to the new plant site, which can be costly both from the standpoint of the physical move and with regard to the additional expense of relocating a family from one place to wholly new surroundings. Older and less flexible work forces often choose to accept early retirement, quit, or not to relocate.

Training a new work force can also be costly. Bringing a new employee to the required level of proficiency requires time, and does not always succeed. It is therefore preferred to have access to a work force with developed skills that can be readily converted to suit the special requirements of a new business. The new plant site needs to be accessible to a

sufficiently large work force that the incremental new requirements do not cause a shortage to develop that could raise the wage rates above prevailing rates. Usually the factors that make a site desirable also include incentives that have already attracted a skilled work force that can be hired away from other, similar employers. It is advisable to have a labor survey conducted well in advance of site selection in order to avoid surprises. This can be done by hiring local recruiters, by running blind advertisements in local newspapers for key positions that are to be filled first, or by using consultants.

12.4 Living Conditions

Quality-of-life considerations, which are considered important in attracting personnel to relocate, are necessary: living conditions, i.e., house availability, costs, and safety; schools, i.e., quality, class size, and distance; healthcare, i.e., availability of a major hospital and local doctors; recreational facilities, i.e., types and proximity; cultural, i.e., musical plays, and movie theaters; and sports, i.e., Major or Minor League teams.

Chapter Thirteen

13.0 ECONOMIC ANALYSIS

13.1 Introduction

For any industrial plant to be put into operation, huge sum of money must have been invested in it to purchase and install the necessary machinery and equipment. Land and service facilities must be obtained and the plant must be erected complete with the piping, controls, and service. In addition to all these cost, it is necessary to have money available for the payment of expenses involved in plant operation (Sinnott, 2003).

The capital requires to have the necessary manufacturing and plant facilities is called the fixed capital investment. That which is necessary for operation of the plant is termed the working capital. The sum of the fixed investment and the working capital is the total capital equipment (TCI)

13.2 Accuracy and Purpose of Capital Cost Estimates

The accuracy of an estimate depends on the amount of design detail available: the accuracy of the cost data available; and the time spend on preparing the estimate. In the early stages of a project only an appropriate estimate will be required, and justified, by the amount of information by then developed (Sinnott, 2003).

13.3 Fixed and Working Capital

13.3.1 Fixed Capital

Fixed capital is the total cost of the plant ready for start-up. It is the cost paid to the contractors.

It includes the cost of:

1. Design, and other engineering and construction supervision.
2. All items of equipment and their installation.
3. All piping, instrumentation and control systems.
4. Buildings and structures.

5. Auxiliary facilities, such as utilities, land and civil engineering work.

It is a once-only cost that is not recovered at the end of the project life, other than the scrap value.

13.3.2 Working Capital

Working capital is the additional investments needed, over and above the fixed capital, to start the plant and operate it to the point when income is earned. It includes the cost of:

1. Start-up.
2. Initial catalyst charges.
3. Raw materials and intermediates in the process.
4. Finished product inventories.
5. Funds to cover outstanding accounts from customers.

Most of the working capital is recovered at the end of the project. The total investment needed for a project is the sum of the fixed and working capital. Working capital can vary from as low as 5 per cent of the fixed capital for a simple, single-product, process with little or no finished product storage; to as high as 30 per cent for a process producing a diverse range of product grades for a sophisticated market, such as synthetic fibres.

13.4 Operating Costs

An estimate of the operating costs, the cost of producing the product, is needed to judge the viability of a project, and to make choices between alternative processing possible schemes. These costs can be estimated from the flow-sheet, which gives the raw material and service requirements, and the capital cost estimate. The cost of producing a chemical product will include the items listed below. They are divided into two groups:

1. Fixed operating costs: costs that do not vary with production rate. These are the bills that have to be paid whatever the quantity produced.

2. Variable operating costs: costs that are dependent on the amount of product produced.

13.4.1 Fixed Operating Costs

1. Maintenance (labour and materials).

2. Operating labour.

3. Laboratory costs.

4. Supervision.

5. Plant overheads.

6. Capital charges.

7. Rates (and any other local taxes).

8. Insurance.

9. License fees and royalty payments.

13.4.2 Variable Operating Costs

1. Raw materials.

2. Miscellaneous operating materials.

3. Utilities (services).

4. Shipping and packaging.

The costs listed above are the direct costs of producing the product at the plant site. In addition to these costs the site will have to carry its share of the company's general operating expenses. These will include:

1. General overheads.

2. Research and development costs.

3. Sales expense.

4. Reserves.

13.5 Computer-Aided Algorithm for Costing

Most large manufacturing and contracting organisations use computer programs to aid in the preparation of cost estimates and in process evaluation. Many have developed their own programs, using cost data available from company records to ensure that the estimates are reliable.

Consequently, in order to the cost analysis of this design project, a computer software known as MathCAD will be used to prepare the cost analysis algorithm. In addition, where necessary, Excel package will also be used.

13.6 Economic Analysis Calculations

An approximate estimate of the capital project can be obtained from knowledge of the cost of earlier projects using the same manufacturing process.

The capital cost of a project is related to capacity by the equation,

$$C_2 = C_1 \left(\frac{S_2}{S_1} \right)^N$$

Where

C_1 = “capital cost of the project with capacity S_1 ”

C_2 = “Capital cost of the project with capacity S_2 ”

N = “index number”

S_1 = “Capacity in year 1”

S_2 = “Capacity in year 2”

13.6.1 Total Capital Investment: Using the Factorial Method

Procedure

Prepare material and energy balances, draw up preliminary flow-sheets,

1. Size major equipment items and select materials of construction.
2. Estimate the purchase cost of the major equipment items. Use Figures from general literature.
3. Calculate the total physical plant cost (PPC), using the factors given in general literature
4. Calculate the indirect costs from the direct costs using the factors given in general literature

$$PPC = PCE (1+f_1 + \dots+f_9)$$

5. The direct plus indirect costs give the total fixed capital.
6. Estimate the working capital as a percentage of the fixed capital; 10 to 20 per cent.
7. Add the fixed and working capital to get the total investment required.

13.6.2 Estimation of Purchase Equipment Costs

Preflash

$$\text{Volume} = 36.30 \text{ m}^3$$

$$\text{Diameter} = 1.98 \text{ m}$$

$$\text{Height} = 10.90 \text{ m}$$

$$\text{Pressure} = 517.1 \text{ kPa (5.171 bar)}$$

$$\text{Purchased cost} = \text{barecost} \times \text{material factor} \times \text{Pressure Factor}$$

Where:

$$\text{Material factor} = 1 \text{ (Carbon steel)}$$

$$\text{Pressure factor} = 1.1$$

$$\text{Bare cost} = \$ 31,500$$

$$\begin{aligned} \therefore \text{Purchase cost of Preflash} &= 31,500 \times 1 \times 1.1 \\ &= \$ 33,500 \end{aligned}$$

Furnace

$$C_e = CS^n$$

Where:

C_e = Purchase equipment cost

S = Characteristic size parameter (16361.11 kW)

C = Cost constant (\$ 560)

n = index of equipment type (0.77)

$$\begin{aligned}\therefore C_e &= 560 \times 16361.11^{0.77} \\ &= \$ 498,720.20\end{aligned}$$

Distillation Column cost

Distillation C Cost = Bare vessel cost × material factor × pressure factor

Bare vessel cost \$ 41,000; material factor = 1; pressure Factor =1

$$\begin{aligned}\therefore \text{Bare waste} &= 41,000 \times 1.0 \times 1.0 \\ &= \$ 41,000\end{aligned}$$

Tray Cost = \$ 3,600

For 19 trays = 19 x \$ 3,600

Total cost of column = 41,000+68,400 =\$ 139,400

Total purchase cost of major equipment items (PCE)

Table 13.1: Equipment cost price

Equipment	Cost Price (\$)	Cost Price (₹)
Distillation Column	109,400	16,410,000
Preflash Column	33,500	5,025,000
Furnace	498,720.20	74,808,030
Steam Generator	75,000	11,250,000
Mixer	34,000	5,100,000
Total	750,620	112,593,000

Estimation of Fixed Capital Cost

The fixed capital investment of this design project will be estimated from the capital investment of the plant constructed

Equipment cost	0.4
Piping	0.7
Instrumentation	0.20
Electricals	0.1
Buildings	0.15
Utilities	0.50
Ancillary Equipments	0.15

Total Physical plant cost (PPC) = $750,620 (0.4+0.7+0.2+0.1+0.15+0.5+0.15)$

Design and Engineering 0.3

Contingencies 0.1

Fixed capital Investment = $PPC (1+0.3+0.1) = 1, 651,364.44 (1+0.1+0.3)$

= \$2, 131,190.2

Working capital investment

The working capital investment cost (WCIC) was found to be 10-20 % of the fixed capital investment (FCI). Now taking the working capital investment cost (WCIC) is 11 % of the fixed investment (FCI), then

$$\begin{aligned} \text{Working Capital, all 11 \% of fixed capital} \\ = 0.11 \times 2,131,190.2 = \$ 234,430.92 \end{aligned}$$

Total capital investment

The total capital investment is given as the sum of the fixed capital and working capital investments. That is, denoting the total capital investment by TCI

$$\text{TCI} = \text{FCI} + \text{WCIC}$$

$$\begin{aligned} \text{Total Capital Investment} &= \text{Fixed Capital and Working Capital} \\ &= \$ 2,131,190.2 + \$ 234,430.92 \\ &= \$ 2,365,620.92 \\ &= 3,540,843,138 \text{ Naira} \end{aligned}$$

13.6.3 Estimation of Total Production Cost (TPC)

13.6.3.1 Fixed charges

Depreciation

The depreciation (Dep) was found to be 10 % of the fixed capital investment (FCI). Now making the depreciation (Dep) to be 10 % of the fixed capital investment (FCI), then,

$$\begin{aligned} \text{Dep} &= 10 \% \times \text{FCI} \\ \text{Dep} &= 0.1 \times \$ 2,131,190.2 \\ &= \$ 213,119.02 \\ &= 31,967,853 \text{ Naira} \end{aligned}$$

Local taxes

The local taxes (LT) were found to be 1-2% of the fixed capital investment (FCI). Now taking the local taxes (LT) to be 1.5% of the fixed capital investment (FCI), then

$$LT = 1.5 \% \times FCI$$

$$= 0.015 \times \$ 2, 131,190.2$$

$$=\$ 31,967.85$$

Insurance

The insurance (Ins) was found to be 0.4-1% of the fixed capital investment (FCI), then

$$Ins = 0.7\% \times FCI$$

$$= 0.007 \times \$ 2, 131,190.2$$

$$= \$ 14,918.33$$

Rent

The rent (Rent) was found to be 8-12% of the fixed capital investment (FCI). Taking the rent (Rent) to be 9% of the fixed capital investment (FCI), then

$$Rent = 9 \% \times FCI$$

$$= 0.009 \times \$ 2, 131,190.2$$

$$= \$ 19,180.12$$

Therefore, the total production costs which is the sum of the depreciation, local taxes, insurances and rent is equal to

$$TPC = Dep + LT + Ins + Rent$$

$$TPC = 213,119.02 + 31,967.85 + 14,918.33 + 19,180.12$$

$$TPC = \$ 279,185.32$$

$$\text{TPC} = \text{N}41,877,789$$

Direct production costs

Raw material

The cost of raw material (RMC) was found to be 10-50% of the total production cost (TPC). Now, taking the cost of raw material (RMC) is 45% of the total production cost (TPC), then

$$\text{RMC} = 45\% \times \text{TPC}$$

$$= 0.45 \times \text{N}41,877,789$$

$$= \text{N}18,845,009.1$$

Operating labour

The cost of operating labour (OLC) was found to be 10-20 % of the total production cost (TPC). Now, taking the cost of operating labour (OLC) to be 13 % of the total production cost (TPC), then

$$\text{OLC} = 13\% \times \text{TPC}$$

$$\text{OLC} = 0.13 \times \text{N}41,877,789$$

$$\text{OLC} = \text{N}5,444,112.57$$

Direct supervisory and electric labour cost

The direct supervisory and electric labour cost (DSEL) was found to be 10-25 % of the operating labour cost (OLC). Now taking the direct supervisory and electric labour cost (DSEL) is 13 % of the operating labour cost (OLC), then

$$\text{DSEL} = 13\% \times \text{OLC}$$

$$\text{DSEL} = 0.13 \times \text{N}5,444,112.57$$

$$\text{DSEL} = \text{N}707,734.63$$

Cost of utilities

The cost of utilities (UC) was found to be 10-20 % of the total production cost (TPC).

Now taking the cost of utilities (UC) is 13 % of the total production cost (TPC), then

$$UC = 13 \% \times TPC$$

$$UC = 0.13 \times 41,877,789$$

$$UC = 5,444,112.57 \text{ Naira}$$

Maintenance cost

The maintenance cost (MC) was found to be 2-10 % of the fixed capital investment (FCI).

Now, taking the maintenance (MC) to be 5 % of the fixed capital investment (FCI),

Then

$$MC = 5 \% \times \$ 2,131,190.2$$

$$MC = 0.05 \times \$ 2,131,190.2$$

$$MC = \$ 106,559.51$$

$$MC = 15,983,926.5 \text{ Naira}$$

Operating supplies cost

The operating supply cost (OSC) is found to be 10-20 % of the maintenance cost (MC).

Now, taking the operating supply cost (OSC) to be 11 % of the maintenance cost (MC),

then

$$OSC = 11 \% \times MC$$

$$OSC = 0.11 \times \$ 106,559.51$$

$$OSC = \$ 11,721.55$$

$$OSC = 1,758,232.5 \text{ Naira}$$

Laboratory charges

The laboratory charges (Lab C) was found to be 10-20 % of the Operating cost (OLC).
Now, assuming that the laboratory charges (Lab) is 13 % of the operating labour cost (OLC), then

$$\text{Lab C} = 13 \% \times \text{OLC}$$

$$\text{Lab C} = 0.13 \times 5,444,112.57 = 707,734.63 \text{ Naira}$$

Patent and royalties

The patent and royalties (PR) was found to be 2-6 % of the total production cost (TPC).
Now, taking the patent and royalties (PR) to be 3 % of the total production cost

$$\text{PR} = 3 \% \times \text{TPC}$$

$$\text{PR} = 0.03 \times 41,877,789 = 125,633.67 \text{ Naira}$$

Plant overhead cost

The plant overhead cost (OVC) was found to be 50-70 % of the sum of operating labour (OLC), operating supplies (OSC) and maintenance cost (MC). Now, taking the plant overhead cost (OVC) is 53 % of the sum of operating labour (OLC), operating supplies (OSC) and maintenance cost (MC), then

$$\text{OVC} = 53 \% \times (\text{OLC} + \text{OSC} + \text{MC})$$

$$\text{OVC} = 0.53 (5,444,112.57 + 1,758,232.5 + 15,983,926.5)$$

$$\text{OVC} = 23,186,71.57 \text{ Naira}$$

13.6.4 General Expenses

13.6.4.1 Administrative cost

The administrative cost (AC) was found to be 10-60% of the operating labour cost (OLC), then

$$AC = 55 \% \times OLC$$

$$AC = 0.55 \times N 5,444,112.57 = N 2,994,261.91$$

13.6.4.2 Distribution and selling price

The distribution and selling price (DSP) was found to be 2-30 % of the total production cost (TPC). Now, taking the distribution and selling price (DSP) to be 20 % of the total production cost (TPC), then

$$DSP = 20 \% \times TPC$$

$$DSP = 0.2 \times 41,877,789 = 837,557.80 \text{ Naira}$$

So, the general expense (GE) is given as the sum of the administrative cost, distribution and selling price and research and development cost. That is,

$$GE = AC + DSP + RDC$$

$$GE = 2,994,261.91 + 837,557.80 + 837,557.80$$

$$GE = 4,669,377.51 \text{ Naira}$$

13.6.4.3 Total product cost

The total product cost (TProdC) is given as the sum of the maintenance cost and the general expenses. That is

$$TProdC = MC + GE$$

$$TProdC = 15,983,926.5 + 4,669,377.51$$

$$TProdC = 20,653,304.01 \text{ Naira}$$

Further, the total expenditure, which is given as the sum of raw material, factory labour, depreciation and overhead, is equal to

$$TE = RMC + OLC + Dep + OVC$$

$$TE = 218,845,009.10 + 5,444,112.57 + 31,967,853 + 23,186,271.57$$

$$TE = 279,443,246.24 \text{ Naira}$$

13.6.5 Revenue Expectations

Having known the cost price, the selling price can now be set. Setting the price average to be 90 naira/L, that is,

$$\text{Profit} = \text{Amount per Annum price of products (cut)}$$

$$\text{Profit} = 9,006,330.19 \times 4 \times 90$$

$$\text{Profit} = 810,569,716.94 \frac{\text{Naira}}{\text{annum}}$$

$$TI = \text{Profit}$$

$$TI = 810,569,716.94 \text{ Naira}$$

Gross income (GI) is the difference between the total income and the total expenditure (TE).

That is,

$$GI = TI - TE = 810,569,716.94 - 279,443,246.24$$

$$GI = 531,126,470.70 \text{ Naira}$$

The total income above is the name before tax. Giving that tax is 25 % of the total income,

$$\text{That is, Tax } 25 \% \times GI = 0.25 \times 531,126,470.70 = 132,781,617.67 \text{ Naira}$$

Net profit (NP) is thus obtained as the difference between the gross income (GI) and the tax (Tax). That is,

$$NP = GI - Tax$$

$$NP = 531,126,470.70 - 132,781,617.67$$

$$NP = 708,168,827.60 \text{ Naira}$$

Rate of return (PR) is given as

Rate of return (ROR) is given as

$$\text{Rate of return} = \frac{\text{Net profit}}{\text{Total capital investment}} \times 100.$$

$$\text{Rate of return} = \frac{708,168,827.60}{3,540,843,138} \times 100.$$

$$\text{ROR} = 20 \%$$

So, symbolically,

$$\text{ROR} = \frac{NP}{TCI} \times 100\%$$

13.6.5.1 Cash flow

Cash flow is given as the difference between the amount earned and the amount expended.

That

$$\text{Cash flow} = TI - T_{\text{prodC}}$$

$$\text{Cash flow} = 3,540,843,138 - 41,877,789 = 3,498,965,349 \text{ Naira}$$

13.6.5.2 Payback period

The payback period (PBP) is the time required after the start of the project to pay off the initial investment from income. It is given as the reciprocal of the rate of return (ROR).

That is,

$$PBP = \frac{1}{ROR} \text{ yr}$$

$$PBP = \frac{1}{0.2} \text{ yr}$$

$$PBP = 5 \text{ yrs}$$

13.6.5.3 Discounted cash flow (time value of money)

The discounted cash flow is used in this project to obtain the Net present Value of the cash flow in year n. the net cash flow in each year of the project is brought to its “present worth” at the start of the project by discounting it at some chosen compound interest rate. The formula for the net present worth (NPW) of cash flow in year n is given as

$$NPW = \sum_{i=1}^n \frac{\text{Cash flow}}{(1+r)^n}$$

Now for this project, the discount rate $r = 20\%$ for year 1

Therefore,

NPW

=

$$\sum_{i=1}^n \frac{\text{Cash flow}}{(1+r)^n}$$

$$\sum_{i=1}^n \frac{3,498,965,349}{(1+0.20)^1}$$

$$NPW = 2,915,804,457.50 \text{ Naira}$$

13.6.5.4 Discounted cash flow rate of return

The discounted cash flow rate is given as the interest rate that will make the condition given as

DCF

=

$$\sum_{i=1}^n \frac{\text{Cash flow}}{(1+r)^n} = C$$

Using trial-and-error calculations as suggested by (Sinnott, 2003).

$$R = 54.15 \text{ x } \% \quad n = 2$$

$$\text{DCFRR} = R$$

$$\text{DCFRR} = 54.15 \%$$

$$\text{DCF} = \sum_{i=1}^n \frac{\text{Cash flow}}{(1+r)^n}$$

$$\text{DCF} = \sum_{i=1}^{20} \frac{3,498,965,349}{(1+0.542)^2}$$

$$\text{DCF} = 1,471,536,009.90 \text{ Naira}$$

Chapter Fourteen

14.0 CONCLUSION AND RECOMMENDATION

14.1 Conclusion

The major objective of this design is to develop a cost effective and environmental friendly process that takes into consideration the technological development of a viable economy and market structure. Hence, the design of a plant to refine 4800 m³/day of pre-heated crude has been completed. The completed plant was properly sized and specified after carrying out the detailed material and energy balances around each units and hence the overall plant. Safety and effective waste disposal measures were considered and thus the plant can be said to be environmentally acceptable. From the economic analysis evaluated the plant has a payback period of 5 years and a rate of return on investment of 20 % on a total capital invested amount of 2,365,620.92 USD (3,540,843,138.00 Naira).

14.1 Recommendation to the Industrialist

Based on this design work that has been carried out, the following recommendations are made to the industrialists to be noted during the construction, start-up and operating phases of the work:

1. The safety of workers, equipments and infrastructures should be highly evaluated during the design implementation stage of the design.
2. The recycle design should be considered without further delay, since it is useful in reducing the level of plant effluent, and it also reduces costs.
3. Adequate data and technological parameters should be at the possession of the plant operations at all time to forestall any unwanted accident.
4. Routine turn around plant maintenance should be of paramount importance in the design. An articulate and organised maintenance team should safeguard quick plant shut down and ensure equipment salvage value. This will also take care of schedule, slippage, cost over-run and possible re-work.
5. Personnel should undergo routine training about new work ethic and equipments to improve their knowledge of the plant operation and increase overall plant productivity.

6. Procurement of raw materials and equipments should be based on strict regulation of specification and maximum quality.
7. Plant should not be operated above the design specification to avoid abnormal conditions and explosions.
8. The implementation of this design work must be adequately supervised by the experts.
9. The plant should be sited close to the source of raw materials.
10. Alternative sources of energy should be available at all times to avoid plant failure and possible sources of failure.
11. The water and air around the plant should be monitored regularly to ensure compliance with the Environmental Protection Agency Standards.

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