Sudan University of Science and Technology
College of Petroleum and Mining Engineering and Technology
Transportation and Refining Engineering Department

Basic Design in Addition of New Distillation Column to Obied Refinery

تصميم مبدئي لاضافة برج تقطير لمصفاة الابيض

(Case Study Obied Refinery)

This dissertation is submitted as a partial requirement of B.Sc Degree (Honor) in Petroleum Engineering

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استهلال

بسم الله الرحمن الرحيم

" وَقُلْ رَبِّ زِدْنِي عِلْمًا "   **سورة طه{114}**

صدق الله العظيم

DEDICATION

First of all thanks is for Allah, dedicated to our parents who always devising us, nothing of this could be done without them .

To everyone who inspired our creativity, who always were with us step by step.

To anyone who taught us how to breath in this life.

ACKNOWLEDGEMENT

First and foremost, we would like to thank the supervisor Dr . Monira Mahmoud for her support, outstanding guidance and encouragement. We would also like to express our gratitude appreciation to all doctors and all teachers in Sudan university of science and technology for all the help and guidance they provided through our education and to the other members of our teaching assistant.

ABSTRACT

This project show a new opinion applied in Obeid Refinery which is addition of distillation unit contain distillation column, two heat exchanger and crude oil storage tank. this unit is added mainly to satisfy the demand of diesel in the near area as well as covered the shortage in the demand of the electrical sector. Design of this unit was carried out using ASPEN HYSYS as simulator, however the simulator was applied for the disgn of the distillation column with capacity 25000 bbl/day the methodology consider that the Nile blend is a complex mixture so the simulation results determent the light and heavy key between the products to calculate the minimum number of trays and the minimum reflux to determine the actual number of trays and the dimensions of the column in general. The main outcome of the design work is 41 trays and diameter of 1.5 m and 28.7m high.

المستخلص

يوضح هذا المشروع رأيًا جديدًا تم تطبيقه في مصفاة الابيض وهو إضافة وحدة تقطير تحتوي على عمود تقطير ومبادلين حراريين وخزان للنفط الخام. تمت إضافة هذه الوحدة بشكل أساسي لتلبية الطلب على الديزل في المنطقة القريبة وكذلك تغطية النقص في الطلب على قطاع الكهرباء. تم تصميم هذه الوحدة باستخدام ASPEN HYSYS كمحاكاة ، ولكن تم تطبيق المحاكاة على فصل عمود التقطير بسعة 25000 برميل / يوم ، حيث اعتبرت المنهجية أن مزيج النيل عبارة عن خليط معقد ، لذا فإن نتائج المحاكاة تحدد القطفه الخفيفه والقطفه الثقيله لحساب الحد الأدنى لعدد الصواني والحد الأدنى للراجع لتحديد العدد الفعلي للصواني وأبعاد العمود بشكل عام .النتيجة الرئيسية لأعمال التصميم هي 41 صينية بقطر 1.5 متر وارتفاع 28.7 متر.

Thesis outlines

i. Chapter one contains introduction about background (distillation, simulation) ,case study ,problem statement and project objectives.
ii. Chapter two contains review of literature.
iii. Chapter three contains the summary of the methodology and introduction about (Aspen HYHYS software) in addition to the work steps.
iv. Chapter four contains material balance.
v. Chapter five contains the design calculations.
vi. Chapter sex contains final conclusion and recommendations.

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CHAPTER ONE

1. Introduction:

1.1. Background

1.1.1. Distillation

Distillation is a separation process requires differences to be recognized and utilized. We separate many things by detecting a difference in a physical property, color, size, weight, shapes for example it also requires acting according to such information. Separation by distillation implies a difference in boiling points of two or more materials. The components or compounds making up crude oil are numbered in thousands.

Many of these components have similar physical properties including boiling points that may differ by only a few degrees. Therefore, it is difficult to separate some pure compounds from the complex mixture of components in crude oil by distillation alone.

There are other methods of separation used in a refinery for example, extraction with a solvent, crystallization, and absorption. However distillation is the most common method. Fortunately, rarely need pure compounds and it is often enough to separate groups of compounds from each other by boiling range.

Crude can be separated into gasoline, naphtha, kerosene, diesel oil, gas oil, and other products, by distillation at atmospheric pressure. Distillation is an operation in which vapors rising through fractionating decks in a tower are intimately contacted with liquid descending across the decks so that higher boiling components are condensed, and concentrate at the bottom of the tower while the lighter ones are concentrated at the top or pass overhead. Crude is generally pumped to the unit directly from a storage tank, and it is important that charge tanks be drained completely free from water before charging to the unit. If water is entrained in the charge, it will vaporize in the exchangers and in the heater, and cause a high pressure drop through that equipment.

If a slug of water should be charged to the unit, the quantity of steam generated by its vaporization is so much greater than the quantity of vapor obtained from the same volume of oil, that the decks in the fractionating column could be damaged.

Water expands in volume 1600 times upon vaporization at 100ºC at atmospheric pressure. If crude oil were a final product, it would have just been a low grade fuel struggling to establish itself against coal. If we separate the many compounds in crude oil into groups we find that these groups have characteristics that make them considerably more valuable than the whole crude oil. Some of these groups are products some may be feedstock to other processing units where they are chemically changed into more valuable products. These products, in turn, are usually separated or purified by distillation.

1.1.2. Most common materials in manufacture

1.Carbon steel

Is the most widely used engineering and construction material for industrial applications on a large scale, including marine structures, power plants, transportation, chemical processing and petroleum production and refining.

Carbon steel has high tensile strength and hardness but is significantly more prone to corrosion.

Carbon steel may contain a range of carbon concentrations from 0.01% to 1.5% in a given alloy.

2. **Alloy steel**

Is steel that is alloyed with a variety of elements in total amounts between 1.0% and 50% by weight to improve its mechanical properties. Alloy steels are broken down into two groups: low alloy steels and high alloy steels.

1.1.3. Simulation

Simulation is the process of designing a model of a real system and conducting experiments with this model for the purpose either of understanding the behavior of the system or of evaluating various strategies for the operation of the system.

Over the past decades the use of simulations has been widely accepted in chemical engineering for design and analysis of processes ,The commercial process simulation has proven to be an important tool for plant design and operations and are now considered as state of art for the design, analysis and optimization of chemical processes ,There are several process simulation software packages available in today's market the most widely used simulators are Aspen HYSYS® **,** this program uses in CDU and petroleum industry and its powerful software tool that can be used by engineers to design a plant and process .

2.1. Project objective

Add a fractionation unit with daily capacity 25000 bbls per day in Obied refinery so as to increase the throughput of the refinery from bbls per day to 40000 bbls per day 15000.

This selection is based on that the local production from existing refineries in Sudan cannot satisfy the demand which increases dramatically due to life and economic development So the shortage can be minimized by adding this unit.

The areas around Obied refinery is agricultural area and also rural area which highly need diesel products well as kerosene for domestic use, so the new production can be consumed in the neighboring area ,no need to transport from far away such as like from Khartoum, which can save the fuel consumed in transportation by road.

Also long residue can be used as fuel for power plant in Khartoum since depend at beginning on the refinery old production and by increasing the power station, the demand increases ,so the shortage is covered currently by import which needs hard currency So producing more fuel from this proposed new unit can stop the importation of fuel oil which contributes greatly in economic benefit and stability of power supply.

The crude processed in Obied refinery is Sudanese crude produced in south west of Sudan, The same crude proposed in the design of the added unit.

1.3. Case Study Background

El Obeid Refinery is located to the north of El Obeid, the capital of North Kordofan State in western Sudan. It is registered as a limited liability company, owned by the Sudanese Oil Corporation of the Sudanese Ministry of Energy. The refinery was established in 1996, and the project to establish it took a period not exceeding six months, which is a record number. During that period, the installation of refining units, electrical generating stations, crude oil tanks, lines connecting them to the refining units, shipping ports, crude unloading docks, and shipping ferns were completed. The refinery was inaugurated on July 22, 1996 in a remarkable event. It was one of the days of North Kordofan State that will be remembered by history, emphasizing the embrace and solidarity between the center and the region. The refinery was the rescue gift on its seventh anniversary for the states of Kordofan and Darfour. The refinery started its activity by refining 10,000 barrels per day, then it rose to 15,000 barrels per day, and it refines the incoming crude into four products:

1- Naphtha. 2- Kerosene. 3- Gasoline. 4- The furnace

In addition to the diesel product, it is a mixture of gasoline, kerosene and ferns. Diesel is used in thermal power plants in the states of Greater Darfur, Al-Nuhoud and Kadugli. More about Al-Obeid Refinery

1.2.1.Refinery Products:

1-Naphtha:

It is unprocessed gasoline and is used by mixing it with gasoline or an additive or entering it into the gasoline improvement unit to acquire the specifications of regular gasoline.

2-Kerosene:

It is the first product that drives the economy and is used in the fields of agriculture, transportation, all services, generation and industries.

3-Gasoil

It is the first product that drives the economy and is used in the fields of agriculture, transportation, all services, generation and industries.

4-The furnace

It is used by thermal power plants and steam boilers for sugar, textile, oil and soap factories and other large-scale factories.

1.3.Problem statement

Sufficient local domestic demand for petroleum products

CHAPTER TWO

Literature Review

2.1 Previous work

In2004, an energy analysis of Crude Distillation Unit (CDU) of N'djamena Refinery Company (NRC) Chad Republic was analyzed. In the considered Crude Distillation Unit, the ideal work, lost and shaft work were 2.40E+08, 4.29E+08 and 6.69E+08 Btu/hr. respectively. In addition, the calculated second law efficiency was 35.8% by (M, et al., 2004) . The quality of products of a fractionation column was study considering different design conditions of the column using natural gas condensate as column feed The first design was on a single traditional distillation column whereas the consecutive studies were done on modifying the distillation column to yield the same quality of products keeping the material balance constant. This study includes the details quality variation along with the variation of design. The whole simulation study and analysis was Done on ASPENTM HYSYS 7.1. (Rahmana & kirtaniaa, 2011) a retrofit design methodology and simulation framework for heat-integrated crude oil distillation systems was studies by using HYSYS (Gadalla , et al., 2013).The optimize of gasoline production in all the refineries was studies .The strategy being to first target the CDUs in these refineries. Maximizing the yield of gasoline and its intermediates will directly impact positively on total pool gasoline production using HYSYS program in comparing (Okeke & Osakwe-Akofe, 2003). development of a methodology for the optimization, control and operability of both existing and new production facilities through an integrated environment of different technologies like process simulation, optimization and control systems. Such an integrated environment not only creates opportunities for operational decision making but also serves as training tool for the novice engineers. It enables them to apply engineering expertise to solve challenges unique to the process industries in a safe and virtual environment and also assist them to get familiarize with the existing control systems and to understand the fundamentals of the plant operation was discussed *by* (Yela, 2009).

2.2.Composition of Crude Oils

Crude oil is a complex liquid mixture made up of a vast number of hydrocarbon compounds that consist mainly of carbon and hydrogen in differing proportions. In addition, small amounts of organic compounds containing Sulphur, oxygen, nitrogen and metals such as vanadium, nickel, iron and copper. The crude oil generally classified into :

1. light crude
2. medium crude
3. heavy crude
4. very heavy crude

Table 2.1 Elemental composition of crude oils

Table 1

Element	Composition (wt%)
Carbon	83.0–87.0
Hydrogen	10.0–14.0
Sulphur	0.05–6.0
Nitrogen	0.1–0.2
Oxygen	0.05–2.0

Ni	<120 ppm
V	<1200ppm

2.2.Properties of Crude Oils

- Density, Specific Gravity and API Gravity

Density is defined as the mass of unit volume of a material at a specific temperature. A more useful unit used by the petroleum industry is specific gravity. The API (American Petroleum Institute) gravity is another way to express the relative masses of crude oils.

- Pour Point

The pour point of a crude oil or product is the lowest temperature at which a crude oil is observed to flow under the conditions of the test. Pour point data indicates the amount of long-chain paraffins (petroleum wax) found in a crude oil. Paraffinic crude oils usually have higher wax content than other crude types.

- Viscosity

Viscosity is a measure of a measure of a fluid's resistance to flow, the lower the viscosity of a fluid, the more easily it flows. Like density, viscosity is affected by temperature. As the temperature decreases, viscosity increases. The unit if dynamic viscosity is the millipascal-second (mpa.s).

- Total acid number (TAN)

Total acid number (TAN), defined as the number of milligrams of KOH required neutralizing the acidity of one gram oil, is a commonly accepted criterion for the oil acidity, TAN number greater than 0.5 are classified as highly acidic. High TAN crude oils are commonly encountered in California, Venezuela, the North Sea, Western Africa, India, China and Russia.

2.3. Crude Distillation Unit

2.3.1. Process description

The first process encountered in any conventional Refinery is the atmospheric crude distillation Unit. In this unit the crude oil is distilled to produce distillate streams which will be the basic streams for the refinery product slate. These streams will either be subject to further treating downstream or become feed stock for conversion units that may be in the refinery configuration a schematic flow diagram of an atmospheric crude unit is shown in Figure (2.1) . Crude oil is pumped from storage to be heated by exchange against hot overhead and product side streams in the crude unit. At a preheat temperature of about (200– 250)◦f water is injected into the crude to dissolve salt that is usually present. The mixture enters a de-salter drum usually containing an electrostatic precipitator. The salt water contained in the crude is separated by means of this electrostatic precipitation.

The water phase from the drum is sent to a sour water stripper to be cleaned before disposal to the oily water sewer , it must be understood however that this ‘de-salting’ does not remove the organic chlorides which may be present in the feed. This will be discussed later when dealing with the tower’s overhead system.

The crude oil leaves the de-salter drum and enters a surge drum. Some of the light ends and any entrained water are flashed off in this drum and routed directly to the distillation tower flash zone (they do not pass through to the heater). The crude distillation booster pump takes suction from this drum and delivers the desalted crude under flow control to the fired heater via the remaining heat exchange train. on leaving heat exchanger train, the crude oil is heated in a fired heater to a temperature that will vaporize the distillate products in the crude tower. Some additional heat is added to the crude to vaporize about 5% more than required for the distillate streams.

This is called over flash and is used to ensure good reflux streams in the tower. The heated crude enters the fractionation tower in a lower section called the flash zone.

The heavy portion of the crude leaves the bottom of the tower as fuel or as feed stock for other unit , while the distillate vapors move up the tower counter current to a cooler liquid reflux stream. Heat and mass transfer take place on the fractionating trays contained in this section of the tower above the flash zone. Middle Distillates are drawn out from selected trays (draw-off trays) . The full naphtha vapor is allowed to leave the top of the tower to be condensed and collected in the overhead drum. A Portion of naphtha stream is returned as reflux to control temperature profile

The products are :

1-Naphtha.

2-Kerosene.

3-Light diesel.

4-Heavy diesel.

5-Residue.

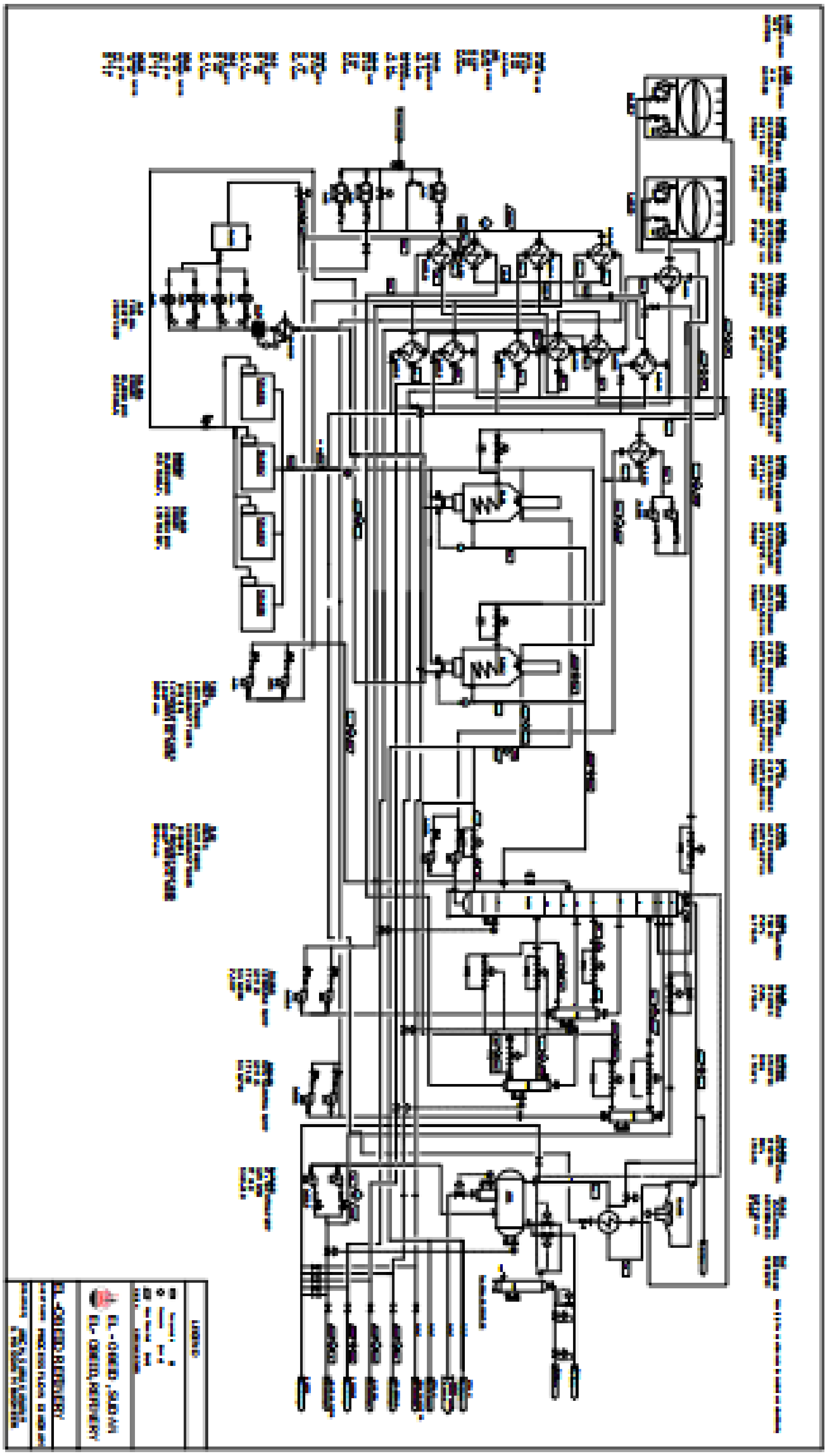


Figure 1 Obeid refinary flow sheet

Figure (2.1) obeid refinery flow sheet

2.3 Oil Characterization in HYSYS

The petroleum characterization method in HYSYS converts the laboratory assay analyses of condensates, crude oils, petroleum cuts, and coal-tar liquids into a series of discrete hypothetical components. These petroleum hypo components provide the basis for the property package to predict the remaining thermodynamic and transport properties necessary for fluid modeling. HYSYS produces a complete set of physical and critical properties for the petroleum hydro-component with a minimal amount of information. However, the more information you can supply about the fluid, the more accurate these properties are, and the better HYSYS predicts the fluid's actual behavior accurate volatility characteristics are vital when representing a petroleum fluid in your process simulation. HYSYS accepts five standard laboratory analytical assay procedures:

1. True boiling point distillation (TBP)
2. ASTM D86 and ASTM D1160 distillations (Separately or Combined)
3. ASTM D2887 simulated distillation.
4. Equilibrium flash vaporization (EFV).
5. Chromatographic analysis.

Chapter Three

Methodology

3.1.CDU Crude Oil Assay

3.1.1Introduction

crude oil assay

For Nile blend, the sample was taken from the inlet of CDU crude tank. The sample was free of any disposal. The Lab set apart wide range of fractions as Naphtha, Kerosene, Diesel, Vacuum distillation, Atmospheric distillation residue and Vacuum distillation residue.

3.1.2. Methods & Instruments:

i. ASTM D2892

(1) The distillation cut from (IBP – 165) ℃ *are* separated at atmospheric pressure.

(2) The distillation cut from (165 – 220) ℃ *are* separated at reduced pressure

(100 mmHg).

(3) The distillation cut from (220 – 350) ℃ *are* separated at reduced pressure (10 mmHg)

ii. ASTM D5236

The distillation cut from (350 – 500) ℃ are separated at reduced pressure (1 mm Hg) by POTSTILL Method.

3.2. Methodology summary for simulation

We will build the (ORC) distillation unit simulation using the following basic steps:

1. Create a unit set.
2. Choose a property package.
3. Select the non-oil components.
4. Characterize the Oil.

5. Create and specify the preheated crude and utility steam streams.
6. Install and define the unit operations in the pre-fractionation Train.
7. Install and define the crude fractionation column.

3.3. HYSYS program

HYSYS is a powerful software tool that can be used by engineers to design plants and processes, optimize production, and enhance decision-making.

3.3.1. Simulation Basis Manager

One of the important concepts upon which HYSYS is based is that of environments. The basis environment allows you to input or access information within the simulation basis manager while the other areas of HYSYS are put on hold.

This helps to maintain peak efficiency by avoiding unnecessary flow sheet calculations once you return to the build environment, all changes that were made in the basis environment take effect at the same time. Conversely, all thermodynamic data is fixed and is not changed as manipulations to the flow sheet take place in the build environment.

3.3.2. The tabs on the Simulation Basis Manager property view Components

Allows access to a component list which is associated with a fluid package. When adding a new component list or editing a current list, the component list property view opens. This property view is designed to simplify adding components to the case like fluid package and hypothetical, oil manager, reactions, component maps user property.

3.3.3. Process simulation procedure:

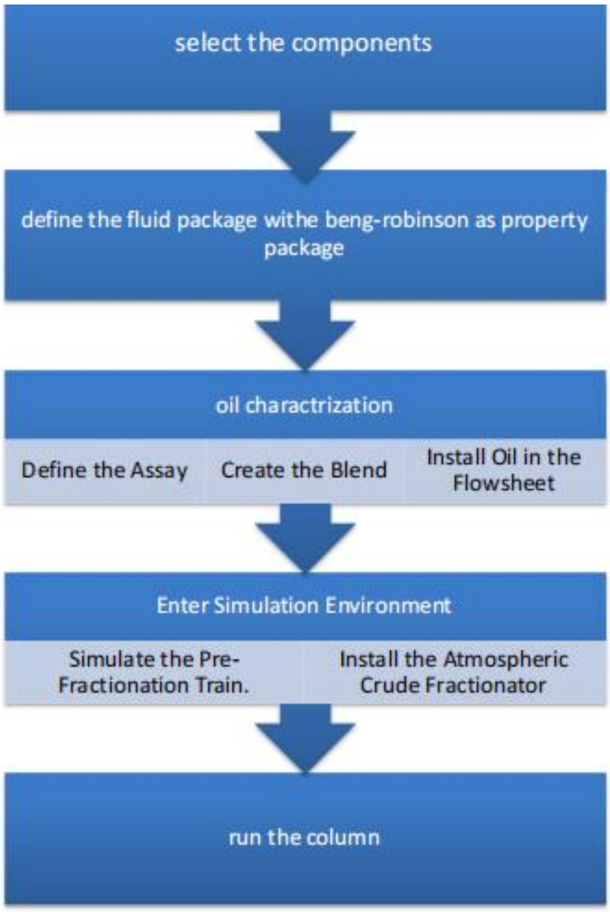


Figure 3.1: process simulation procedure

3.4. Multi component distillation

3.4.1. Two Key Components concept

In the fractionation of multicomponent mixtures, the essential requirement is often the separation of two components. Such components are called the key components and by concentrating attention on these it is possible to simplify the handling of complex mixtures. If a four-component mixture A–B–C–D, in which A is the most volatile and D the least volatile, is to be separated as shown in Figure3.2. then B is the lightest component appearing in the bottoms and is termed the light key component .C is the heaviest component appearing in the distillate and is called the heavy key component. The main purpose of the fractionation is the separation of B from C

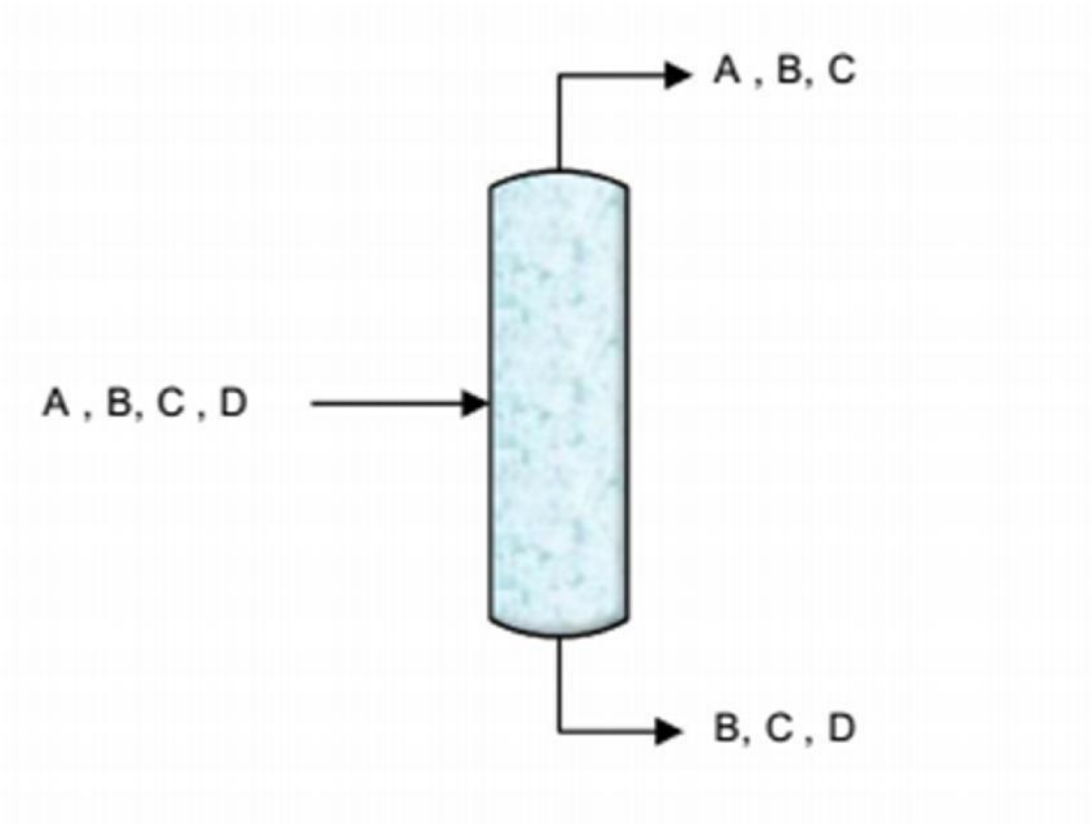


Figure 3.2: distribution of component between top and bottom

3.4.2. Summary designing steps of distillation column:

1. Calculation of Minimum number of stages Nmin
2. Calculation of Minimum Reflux Ratio Rm.
3. Calculation of Actual Reflux Ratio.
4. Calculation of theoretical number of stages.
5. Calculation of actual number of stages.
6. Calculation of diameter of the column.
7. Calculation of the height of the column

3.4.3. Fenske Equation for Minimum Equilibrium Stages

Nmin depends on the degree of separation of the two key components and their mean relative volatility, but is independent of feed-phase condition.

$$N_{min} = \frac{ln\frac{x_{Di}/x_{Bi}}{x_{Dj}/x_{Bj}}}{\ln \alpha_{ij}} - 1 \quad (3.1)$$

Where the mean relative volatility is approximated by

$$\propto_{I,H} = \frac{\propto_{I,K}}{\propto_{H,K}} \quad (3.2)$$

3.4.4. Minimum reflux ratio, using Underwood's method

For feed conditions where the relative volatilities remain constant, UNDERWOOD developed the following two equations from which Rmin may be calculated

$$\frac{\alpha_A x_{fA}}{\alpha_A - \theta} + \frac{\alpha_B x_{fB}}{\alpha_B - \theta} + \frac{\alpha_C x_{fC}}{\alpha_C - \theta} + \cdots = 1 - q \quad (3.3)$$

$$\frac{\alpha_A x_{dA}}{\alpha_A - \theta} + \frac{\alpha_B x_{dB}}{\alpha_B - \theta} + \frac{\alpha_C x_{dC}}{\alpha_C - \theta} + \cdots = R_m + 1 \quad (3.4)$$

where: Xf_A ,X_fB,Xf_C,Xd_A,Xd_B,Xd_C, etc., are the mole fractions of components A, B, C, etc., in the feed and distillate,

A being the light and B the heavy key,

Q: is the ratio of the heat required to vaporize 1 mole of the feed to the molar latent heat of the feed$\propto_A$,$\propto_B$,$\propto_C$, are the volatilities with respect to the least volatile component,

θ: is the root of equation 3.3, which lies between the values of $\propto_A$ and $\propto_B$.

If one component in the system has a relative volatility falling between those of the light and heavy keys, it is necessary to solve for two values of θ. (Richard & Coulson, 2005, pp. 524-525)

3.4.4.1. Relation between reflux ratio and number of plates

GILLILAND has given an empirical relation between the reflux ratio R and the number of plates N, in which only the minimum reflux ratio Rmin and the number of plates at total reflux Nmin are required.

This is shown in Figure3.4, where the group [(N+1) − (Nmin+1)]/(N+2) is plotted against (R−Rmin)/(R+1). (Richardson & Harker, 2002)

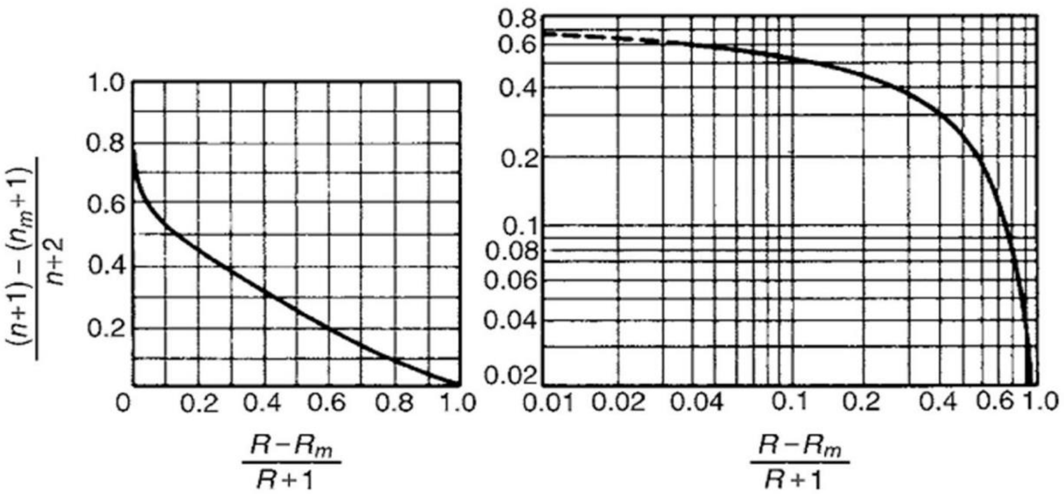


Figure 3.3: Relation between reflux ratio and number of plates.

3.5. Mechanical design of distillation tower

In the mechanical design of the column, tower diameter, tray spacing, and the detailed layout of each tray is considered. Initially, a diameter is established, based on the criterion of absence from liquid entrainment in the vapor stream, and then the weirs and the down comers are designed to handle the required liquid flow. it is then possible to consider the tray geometry in more detail, and, finally, to examine the general operating conditions for the tray and to establish its optimum range of operation. (Richard & Coulson, 2005)

3.5.1. Operating ranges for trays

For a given tray layout there are certain limits for the flows of vapor and liquid within which stable operation is obtained. The range is shown in figure, which relates to a bubble-cap plate. The region of satisfactory operation is bounded by areas where undesirable phenomena occur. (Richard & Coulson, 2005, p. 567)

3.5.2. Plate efficiency

The number of ideal stages required for a desired separation may be calculated by one of the methods discussed previously, although in practice more trays are required than ideal stages. the ratio N/NP of the number of ideal stages n to the number of actual trays n prep resents the overall efficiency e of the column, which

may be 30–100 per cent the main reason for loss in efficiency is that the kinetics for the rate of approach to equilibrium, and the flow pattern on the plate, may not permit an equilibrium between the vapor and liquid to be attained some empirical equations have been developed from which values of efficiency may be calculated, and this approach is of considerable value in giving a general picture of the problem a simple empirical relationship for the overall efficiency, E, of columns handling petroleum hydrocarbons is given by DRICKAMER and BRADFORD who relate efficiency of the column to the average viscosity of the feed by: -Efficiency =13.30-66.8log(μ)………………………….………(3.5)

μL is the viscosity at the mean tower temperature.

(Seader, et al., 2011)

Actual Reflux Ratio

The rule of thumb is:

R = (1.2 ------- 1.5) R ..min ………………………………..(3.6)

R = 1.5 R min

Actual number of stage

Actual number of stages=$\frac{N_{min}}{E}$……………………………………..(3.7)

3.6. Diameter of the tower

3.6.1. Flow parameter

$$F_{LV} = \left(\frac{L_n}{V_n}\right)\left(\frac{\rho_v}{\rho_l}\right)^{0.5} \quad \text{………………………………………………(3.8)}$$

Where:

F_{LV} = liquid Vapor Factor

L_N = mass vapor flow rate

V_N= mass vapor flow rate

ρ_v=vapour density

ρ_l=liquid density

3.6.2. Capacity parameter

Assume tray spacing = 0.5 m

From figure (3.4) flooding velocity, sieve plate

$$V_{nf} = C_{sb}\left(\frac{\sigma}{20}\right)^{0.2} \left(\frac{\rho_l - \rho_v}{\rho_l}\right)^{0.5} \quad \text{………………………… (3.9)}$$

Where:

σ=Surface tension of Mixture

Net column area

$$A_n = \frac{Q_n}{V_n} \quad \text{………………………………………………...(3.10)}$$

Where:

Qv=Volumetric flow rate of vapors

Assume that down comer occupies 15% of cross sectional area (A_C) of column

$A_c = A_n + A_d$ ……………………………………………..………(3.11)

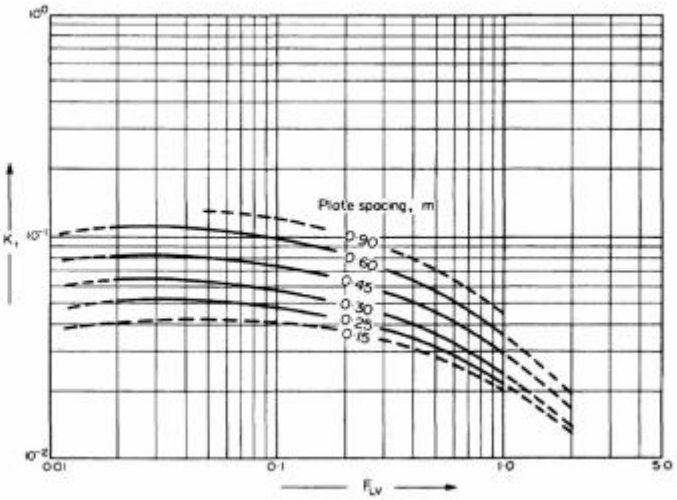


Figure (3.4) Flooding velocity, sieve plate (Richard & Coulson, 2005)

3.7. Height of Distillation Column

$H_T = 2.3 * N_T$ …………………………………………………...(3.12)

Where:

H_T=height of distillation column. (Doglas & James, 1993)

Summary of Design Steps of Heat Exchanger

1-Calculate Energy balance.

2-Calculate LMTD.

3-Calculate Fouling factor.

4-Estemate overall heat transfer coefficient.

5- Calculate heat transfer area.

6-Calculate number of tubes.

7-Calculate number of passes.

1-Energy balance:

Q =m*Cp*ΔT……………………………………………(3.13)

Where:

Q=Heat exchanged (Btu\h)

Cp=specific heat capacity (Btu\Ibm.F)

m=mass flow Ibm\h

ΔT=Temperature difference (F)

2-LMTD:

$$\Delta T_{ln}=\frac{\Delta T_1-\Delta T_2}{Ln(\frac{\Delta T_1}{\Delta T_2})} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3.14)$$

$\Delta T_1=T_{S\,in}-T_{t\,in}$

$\Delta T_2=T_{S\,out}-T_{t\,in}$

3-Fouling factor

$$R=\frac{T_a-T_b}{t_b-t_a} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3.15)$$

$$P=\frac{t_b-t_a}{T_a-t_a}$$

From figure (3.5)

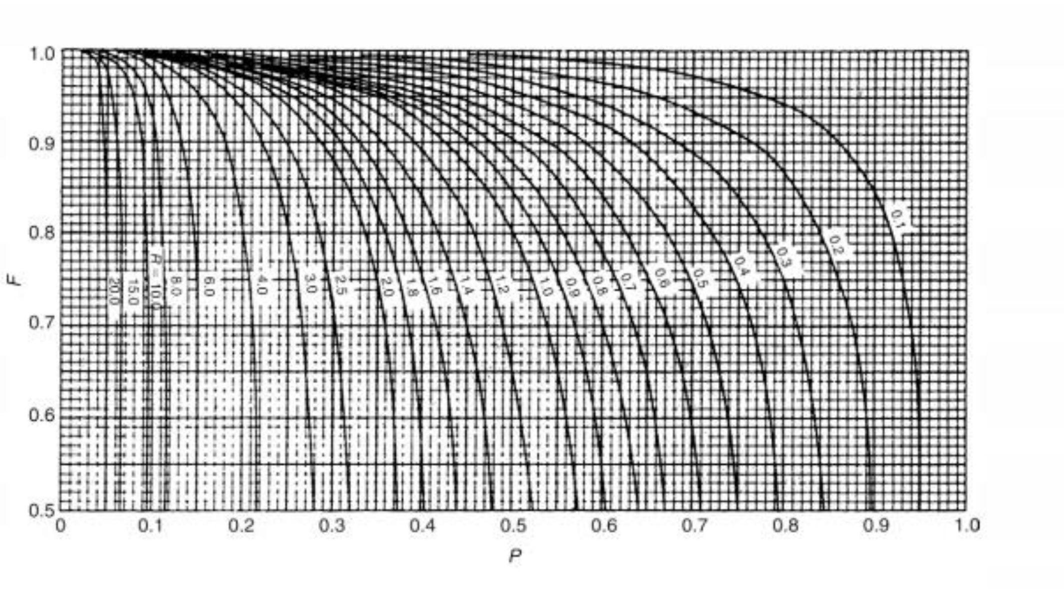


figure (3.5) LMTD correction factor for 1-2 exchangers

3-Estimate U_D

From table (3.2) Typical Values of Overall Heat-Transfer Coefficients in Tubular Heat Exchangers

Table (3.1) Typical Values of Overall Heat-Transfer Coefficients in Tubular Heat Exchangers

Shell side	Tube side	Design *U*	Includes total dirt
Liquid–liquid media			
Aroclor 1248	Jet fuels	100–150	0.0015
Cutback asphalt	Water	10–20	0.01
Demineralized water	Water	300–500	0.001
Ethanol amine (MEA or DEA) 10–25% solutions	Water or DEA, or MEA solutions	140–200	0.003
Fuel oil	Water	15–25	0.007
Fuel oil	Oil	10–15	0.008
Gasoline	Water	60–100	0.003
Heavy oils	Heavy oils	10–40	0.004
Heavy oils	Water	15–50	0.005
Hydrogen-rich reformer stream	Hydrogen-rich reformer stream	90–120	0.002
Kerosene or gas oil	Water	25–50	0.005
Kerosene or gas oil	Oil	20–35	0.005
Kerosene or jet fuels	Trichloroethylene	40–50	0.0015
Jacket water	Water	230–300	0.002
Lube oil (low viscosity)	Water	25–50	0.002
Lube oil (high viscosity)	Water	40–80	0.003
Lube oil	Oil	11–20	0.006
Naphtha	Water	50–70	0.005
Naphtha	Oil	25–35	0.005
Organic solvents	Water	50–150	0.003
Organic solvents	Brine	35–90	0.003
Organic solvents	Organic solvents	20–60	0.002
Tall oil derivatives, vegetable oil, etc.	Water	20–50	0.004
Water	Caustic soda solutions (10–30%)	100–250	0.003
Water	Water	200–250	0.003
Wax distillate	Water	15–25	0.005
Wax distillate	Oil	13–23	0.005
Condensing vapor–liquid media			
Alcohol vapor	Water	100–200	0.002
Asphalt (450°F.)	Dowtherm vapor	40–60	0.006

5- Overall heat transfer coefficient.

$$A = \frac{Q}{U_D * F * \Delta T_{ln}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3.16)$$

Where:

A= area of heat exchanger (ft^2)

6- Number of tubes.

$$n_t = \frac{A}{\pi * D_O * L} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots(3.17)$$

Where:

n_t= number of tubes

L= length of heat exchanger (ft)

7- Shell size

From table (3.2) tube Counts for 3/4-in. OD Tubes on 1-in. Triangular Pitc

Table (3.2) tube Counts for 3/4-in. OD Tubes on 1-in. Triangular Pitc

Shell ID (in.)	TEMA L or M				TEMA P or S				TEMA U		
	Number of passes				Number of passes				Number of passes		
	1	2	4	6	1	2	4	6	2	4	6
8	42	40	26	24	31	26	16	12	32	24	24
10	73	66	52	44	56	48	42	40	52	48	40
12	109	102	88	80	88	78	62	68	84	76	74
13¼	136	128	112	102	121	106	94	88	110	100	98
15¼	183	172	146	148	159	148	132	132	152	140	136
17¼	237	228	208	192	208	198	182	180	206	188	182
19¼	295	282	258	248	258	250	228	220	266	248	234
21¼	361	346	318	320	320	314	290	276	330	316	296
23¼	438	416	382	372	400	384	352	336	400	384	356
25	507	486	448	440	450	442	400	392	472	440	424
27	592	574	536	516	543	530	488	468	554	528	502
29	692	668	632	604	645	618	574	556	648	616	588
31	796	774	732	708	741	716	666	648	744	716	688
33	909	886	836	812	843	826	760	740	852	816	788
35	1023	1002	942	920	950	930	878	856	974	932	908
37	1155	1124	1058	1032	1070	1052	992	968	1092	1056	1008
39	1277	1254	1194	1164	1209	1184	1122	1096	1224	1180	1146
42	1503	1466	1404	1372	1409	1378	1314	1296	1434	1388	1350
45	1726	1690	1622	1588	1635	1608	1536	1504	1652	1604	1560
48	1964	1936	1870	1828	1887	1842	1768	1740	1894	1844	1794
54	2519	2466	2380	2352	2399	2366	2270	2244	2426	2368	2326
60	3095	3058	2954	2928	2981	2940	2832	2800	3006	2944	2884
66	3769	3722	3618	3576							
72	4502	4448	4324	4280							
78	5309	5252	5126	5068							
84	6162	6108	5964	5900							
90	7103	7040	6898	6800							
96	8093	8026	7848	7796							
108	10260	10206	9992	9940							
120	12731	12648	12450	12336							

Chapter 4

Material Balance and Energy Balance

4.1. Material Balance

4.1.1. Mass balance around flash tower

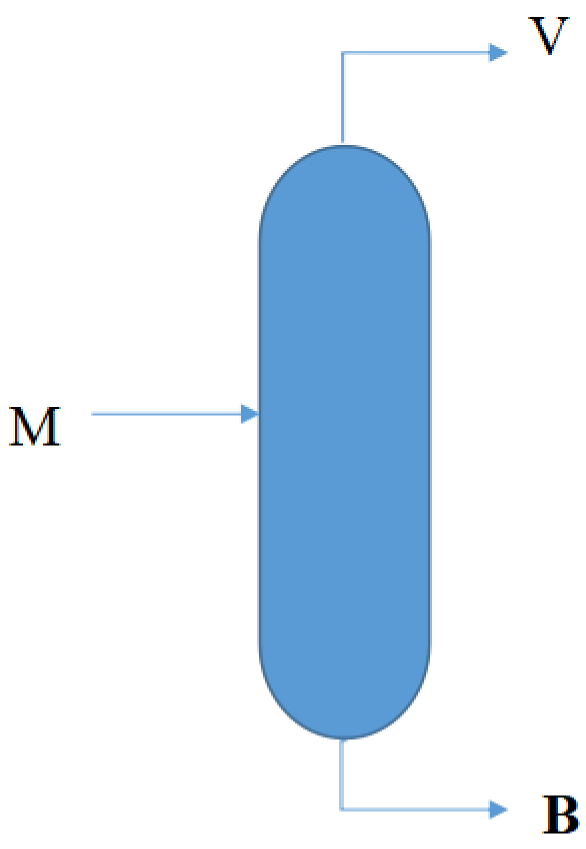


Figure (4.1) Flash tower

Total flow = 173.6 ton\hr

$\frac{V}{M} = 0.00273$

At top

V = 173.6 * 0.00273 = 0.473928 ton\hr

At bottom

B = 173.6*0.9973 = 173.13128 ton\hr

4.1.2. Material balance around distillation column

The equation of material balance for any system:

Input + Generating – Consumption – Output = Accumulation

The mass flow rate

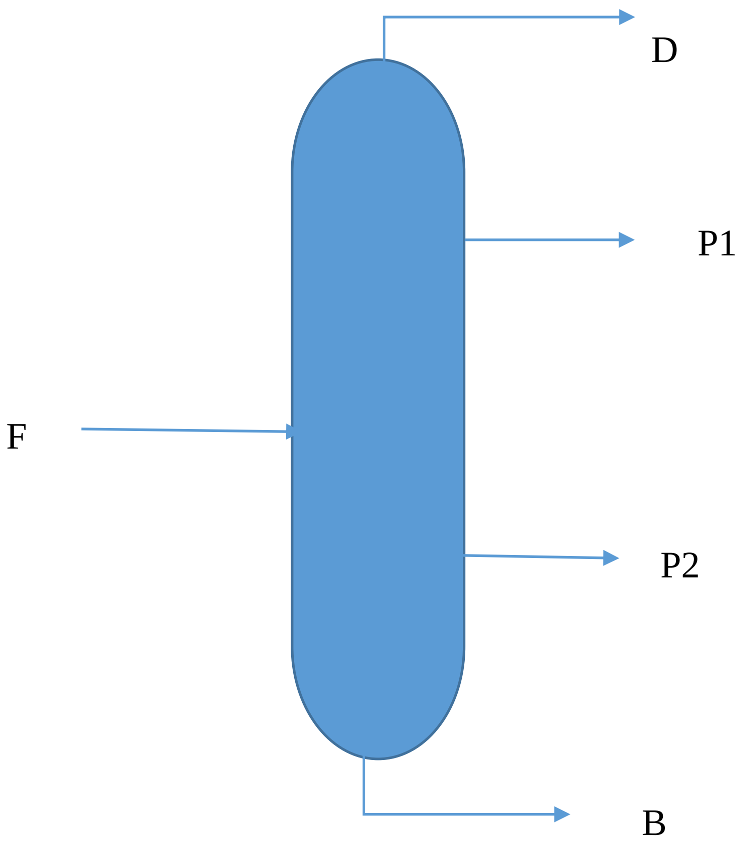


Figure (4.2) Overall Material Balance

At steady state

- Accumulation = 0
- Generation = 0
- Consumption = 0

Then

Input = Output

For overall material balance:

F = D + P1 + P2 + B

F = Feed

D = Naphtha

P1= kerosene

P2= Diesel

B = Residue

Table 4.1 fractions and specific gravity of the product

Component	Concentration in feed	S.G
D	6	0.711
P1	6	0.776
P2	23	0.84
B	65	0.894
Total	100%	

The mass flow rate of

D = 173.6*0.06=10.416 ton\hr = 25000*0.06=1500 bbl\day

The mass flow rate of

P1= 173.6*0.06=10.416 ton\hr = 25000*0.06=1500 bbl\day

The mass flow rate of

P2= 173.6*0.23=39.928 ton\hr = 25000*0.23=5750 bbl\day

The mass flow rate of

B =173.6*0.65=112.85 ton\day = 2500*0.65=16250 bbl\day

Crude = Naphtha +Kerosene + Diesel + Residue

Table 4.2. mass balance around distillation column

Product	Cut mass %	Mass flow Ton\h	Mass flow Kg\h	Volume flow $m^3\backslash h$	Volume flow bbl\day	Density Kg\$m3$
Naphtha	6	10.416	10416	9.9366	1500	711.3
Kerosene	6	10.416	10461	9.9366	1500	776.1
Gasoil	23	39.928	39628	38.0903	5750	810.9
Residue	65	112.85	112850	107.6465	16250	849.1
Crude	100	173.6	173600	165.61	25000	860.1

4.2. Energy analysis

4.2.1. Units operation energy analysis:

rate energy input = rate energy output

Qin = Q out

Q=mCpΔT…………………………………………………………(4.5)

Where

Q = heat quantity or duty in Kj/hr

m=mass flow rate in Kg/hr

Cp= specific heat capacity in KJ/Kg.C

ΔT = temperature change in C°

1. Preheater

173600 kg/h of Nile blend to be heated from 45 C° to 232.2 C° by exchanging with

hot product's stream.

We assume steady state: accumulation = zero

rate energy input = rate energy output

Qin = Qout

Heat load on preheater, Q = 173600*2.3 *(232.2 – 45) = 7.47*10^7kj/hr

2. Furnace duty

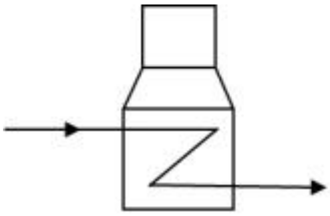


Figure 4.3: furnace input and output streams

We assume steady state: accumulation = zero

rate energy input = rate energy output

Qin = Qout

From HYSYS case

Cp =2.7 KJ/Kg.℃

Duty=145700*2.7*(370 – 232.2)= 5.4 *10^8kj/hr

1. Condenser duty

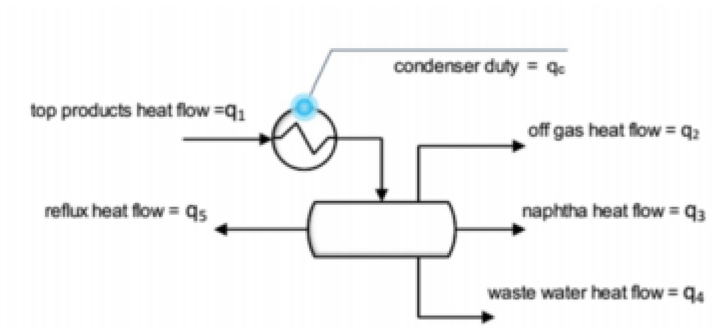


Figure 4.4: Condenser input and output streams

Steady state accumulation =zero

Energy In = energy out

Condenser duty = 3.57*10^8 kj/kg

Chapter Five

Design Calculations

5.1. Design of crude distillation tower

5.1.1. Number of stage and total reflux

1. Minimum number of stage and minimum reflux ratio

Fenske Equation: minimum number of theoretical trays

Minimum number of trays for distillation can be estimated from the following Expression

$$N_{min} = \frac{ln\frac{x_{Di}/x_{Bi}}{x_{Dj}/x_{Bj}}}{\ln \alpha_{ij}} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3.1)$$

$$\alpha_{ij} = \frac{\alpha_{Lk}}{\alpha_{Hk}}$$

$$\alpha_{ij} = 1.74$$

$$N_{min} = \frac{ln\frac{0.0588/0.0012}{0.00046/0.02254}}{\ln 1.74} - 1$$

$= 13.0528$ trays

Minimum reflux ratio

$$\sum \frac{\boldsymbol{\alpha_i x_{i,f}}}{\boldsymbol{\alpha_i - \theta}} = \mathbf{1} - \boldsymbol{q} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \mathbf{(3.2)}$$

$$\sum \frac{\boldsymbol{\alpha_i x_{i,d}}}{\boldsymbol{\alpha_i - \theta}} = \boldsymbol{R_m} + \mathbf{1} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \mathbf{(3.3)}$$

1. **Naphtha and Kerosene**

   From equation (3.2)

   θ = 1.04316

   From equation (3.3)

   R_m+1 = 1.24687

   R_m= 0.24687

2.kerosene and Diesel

From equation (3.2)

θ = 0.7386

From equation (3.3)

R_m+1 = 1.29665

R_m= 0.29665

3.Diesel and residue

From equation (3.2)

θ =3.3723

From equation (3.3)

R_m+1 = 1.05656

R_m= 0.05656

Then:

$R_m = \sum R_m$ = (0.24687)+(0.29665)+(0.05656) = 0.6000

2. Actual reflux ratio

The rule of thumb is:

R = (1.2 ……1.5) R_m

R = 1.2 * R_m

R = 0.840098

$$X = \frac{R-R_m}{R+1} = \frac{0.840098-0.60007}{0.840098+1} = 0.130443$$

$$Y = (1- X^{1/3}) = (1- 0.130443^{1/3}) = 0.4928$$

3. Theoretical number of trays

The Gilliland Correlation (1940) used to calculate the number of stages given reflux ratio and N_{min}

$$N = \frac{N_{min}+Y}{1-Y} = \frac{13.0528 + 0.4928}{1-0.4928}$$

= 26.7066 trays

4. Actual number of stages

Actual number of stages = $\frac{N_{th}}{E}$

The overall tray efficiency is defined as the ratio of number of ideal trays by number of tray required

$$E = 0.17 - 0.616 \log(\mu)$$

Feed viscosity at average temperature $\mu = 0.18$

E = 13.3– 66.8 * log (0.18) = 0.63

$$N = \frac{26.7066}{0.63} = 41 \text{ trays}$$

5.1.2 Determination of Column Diameter

1. Flow parameter

$$F_{LV} = \left(\frac{L_n}{V_n}\right)\left(\frac{\rho_v}{\rho_l}\right)^{0.5}$$

$$F_{LV} = \frac{1.089e+005}{3.679e+004} * \left(\frac{8.697}{662.9}\right)^{0.5}$$

$$= 0.33904$$

2. Capacity parameter

Assume tray spacing = 0.5 m

From figure (3.4) flooding velocity , sieve plate

$C_{SB} =$ 0.4 m\s

$$V_{nf} = C_{sb}\left(\frac{\sigma}{20}\right)^{0.2}\left(\frac{\rho_l - \rho_v}{\rho_l}\right)^{0.5}$$

$$V_{nf} = 0.4\left(\frac{0.009678}{20}\right)^{0.2}\left(\frac{662.9 - 8.697}{8.697}\right)^{0.5}$$

$$= 0.75367 \text{ m\textbackslash s}$$

Assume 90% of flooding then

$V_n = 0.9 V_{nf}$

So , actual vapor velocity ,

$V_n = 0.678303$ m\s

3. Diameter of tower

Net column area

$$A_n = \frac{Q_n}{V_n}$$

$$Q_V = \frac{V_N}{3600 * \rho_V}$$

$Q_V = \frac{3.679e004}{3600*8.697}$

$= 1.17505\ m^3\backslash s$

Now, net area

$A_n = \frac{1.17505}{0.678309}$ = 1.7323 m^2

Assume that down comer occupies 15% of cross sectional area (A_C) of column

$$A_c = A_n + A_d$$

$A_c = A_n + 0.15A_c$

$A_c = \frac{A_n}{0.85}$

$A_c = \frac{1.7323}{0.85}$ = 2.038 m^2

$A_c = \frac{\pi}{4} * D^2$

D = $\sqrt{\frac{4*A_c}{\pi}}$

D = 1.6 m

5.1.3. Height of distillation column

$$H_T = 2.3 * N_T$$

$H_T = 2.3 * 41$ = 94.3 ft = 28.7 m

5.2. Design of heat exchanger:

5.2.1. Energy balance.

Q =m*Cp*ΔT

Q = 10142*0.5698*143.6

=829851.7 Btu/hr

5.2.2. LMTD.

$\Delta T_{ln}=\frac{\Delta T_1-\Delta T_2}{Ln(\frac{\Delta T_1}{\Delta T_2})}$

$\Delta T_1=T_{S\ in}$- $T_{t\ in}$

ΔT_1= 260.7-129.7 =

131 °F

$\Delta T_2 = T_{S\ out} - T_{t\ in}$

$\Delta T_2 = 117.1 - 70$ = 47.1 °F

$\Delta T_{ln} = \frac{\mathbf{131-47.1}}{\ln\left(\frac{\mathbf{131}}{\mathbf{47.1}}\right)}$ =

82.02 °F

5.2.3. LMTD correction factor.

R = $\frac{T_a-T_b}{t_b-t_a}$

R = $\frac{260.7-117.1}{129.7-70}$ =

2.4

P = $\frac{t_b-t_a}{T_a-t_a} =$

P = $\frac{129.7-70}{260-70}$ = 0.31

From figure (3.5), for a 1-2 exchanger

F = 0.88

5.2.4 Estimate $\boldsymbol{U_D}$.

From table (3.1) for naphtha / crude oil exchanger , it is found that the overall heat transfer coefficient (25 ≤ U_D ≤ 35) Btu/hr.ft^2°F . A value near the middle of the range is selected : $U_D = 30$ Btu /h.ft^2.°F

5.2.5. Calculate heat-transfer area and number of tube

A = $\frac{Q}{U_D*F*\Delta T_{ln}}$ =

981.7 ft^2

$n_t = \frac{A}{\pi*D_O*L}$ =

250 tube

5.2.6 Number of tube passes

Re = $\frac{4*m*(^{n_p}/_{n_t})}{\pi*D_i*\mu}$ =

Re =$\frac{4*28058*(^{n_p}/_{250})}{\pi*(^{0.74}/_{12})*0.72}$ =

3218.4 n_p

We want Re ≥ 10^4 and an even number of passes.therefore , take $n_p = 6$.

Checking the fluid velocity.

V = $\frac{m^\circ*(n_p/n_t)}{\rho*\pi*D_i^2/4}$ =

1.1 Ft\sec

The velocity at the high end of the recommended range but still acceptable therefore, six tube passes will be used.

5.2.7. Determine shell size

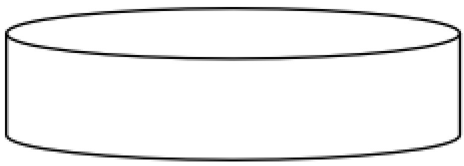
From the tube-count table for $^3/_4$ in . tube on 1 in. square pitch table (3.2) , with Six tube passes and type S head , the listing closest to 250 is 248 tube ,so the Shell diameter is 19$\frac{1}{4}$.

5.3. Design of tank

Assume that the ratio between the length and diameter is

D =3 L

D: diameter



L: length

Q = 4000 m^3/day

Volume/ day = 4000 m^3

Storage duration = 15 days

Volume / 15 day =15*4000= 60000 m^3/day

$$V = \frac{\pi}{4} * D^2 * L$$

$$60000 = \frac{\pi}{4} * (3L)^2 * L$$

$$L = \sqrt[3]{\frac{4*60000}{9\pi}}$$ =20.3 m

D =3*20.3= 60.6 m

Chapter Six

Results and discussion

6.1. Result from

6.1.1. Design of crude distillation tower

Table (6.1) design result (distillation)

No. of tray	41	Tray thickness	0.5m
Pressure	10psi	Reflux ratio	0.60007
Height of column	28.7m	Tray spacing	0.5m
Diameter of Column	1.5m		

6.1.2. Design of heat exchanger

Table (6.2) design result (heat exchanger)

Energy balance.	829851.7 Btu/hr	Number of tube passes	250
LMTD .	82.02 °F	ID	3\4 in
Correction facto	0.88	OD	1 in
Heat-transfer area	981.7 ft^2	Shell diameter	19 1\4 in

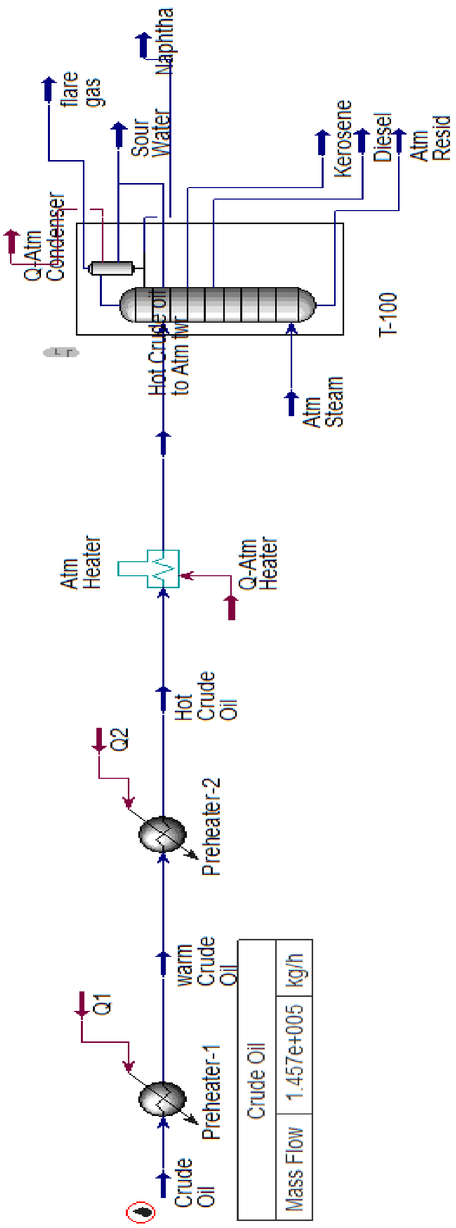


figure (6.1) HYSYS flow sheet 1

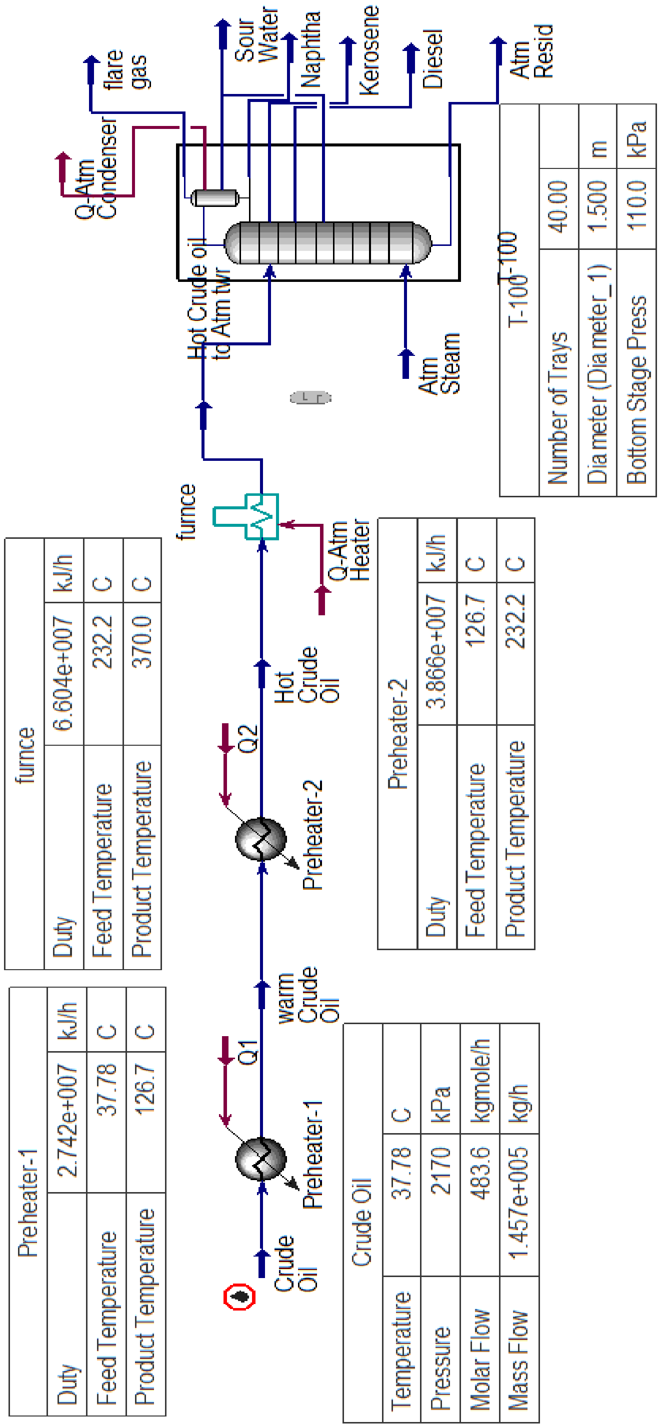


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figure (6.2) HYSYS flow sheet 2

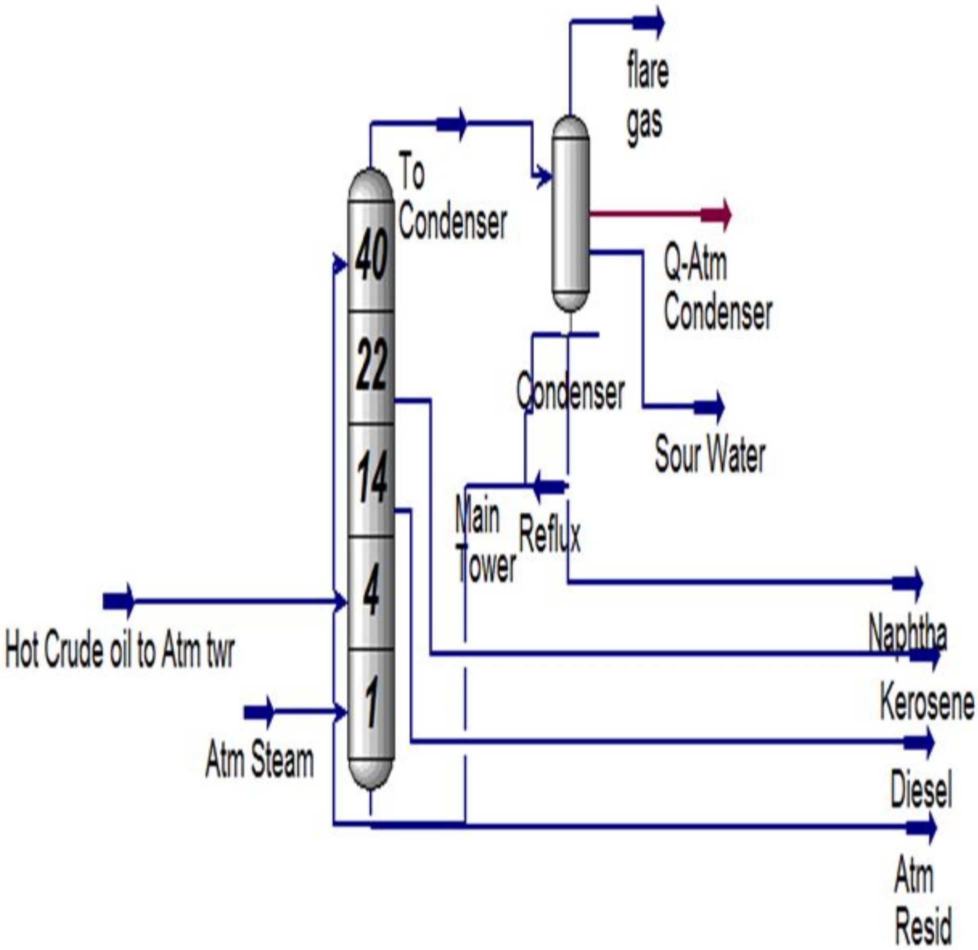

figure (6.3) distillation column from simulation

6.2. curve result

6.2.1. Nile Blend properties

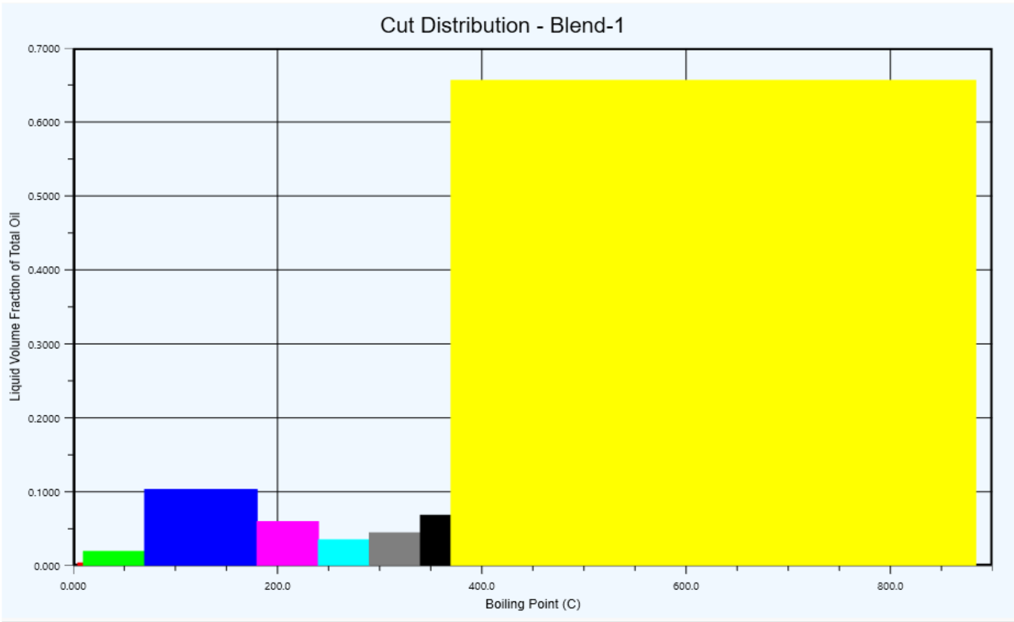

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Figure (6.4) TBP Distillation

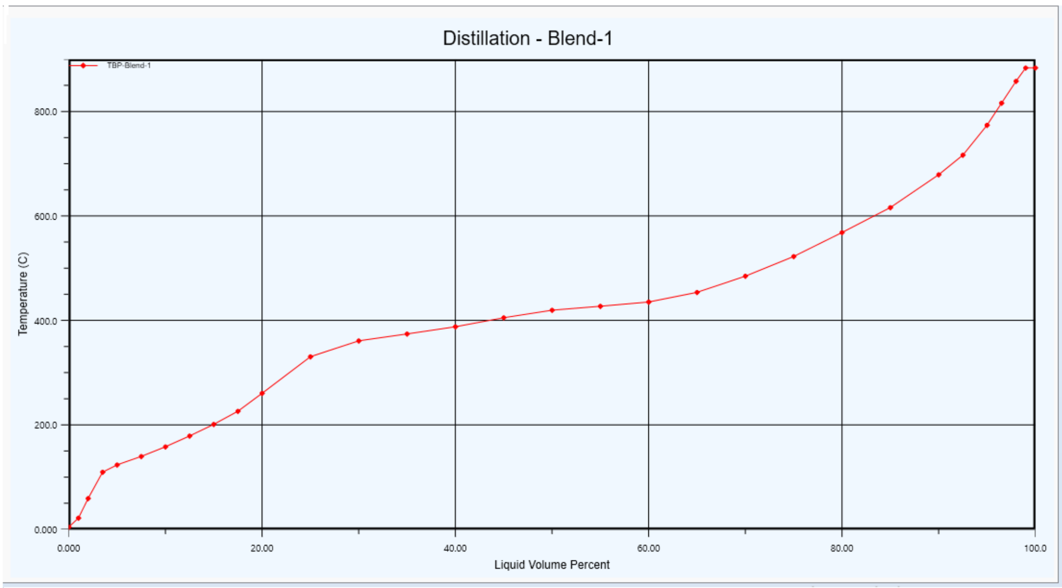

Figure (6.5) Cut distribution

6.2.2. column profile

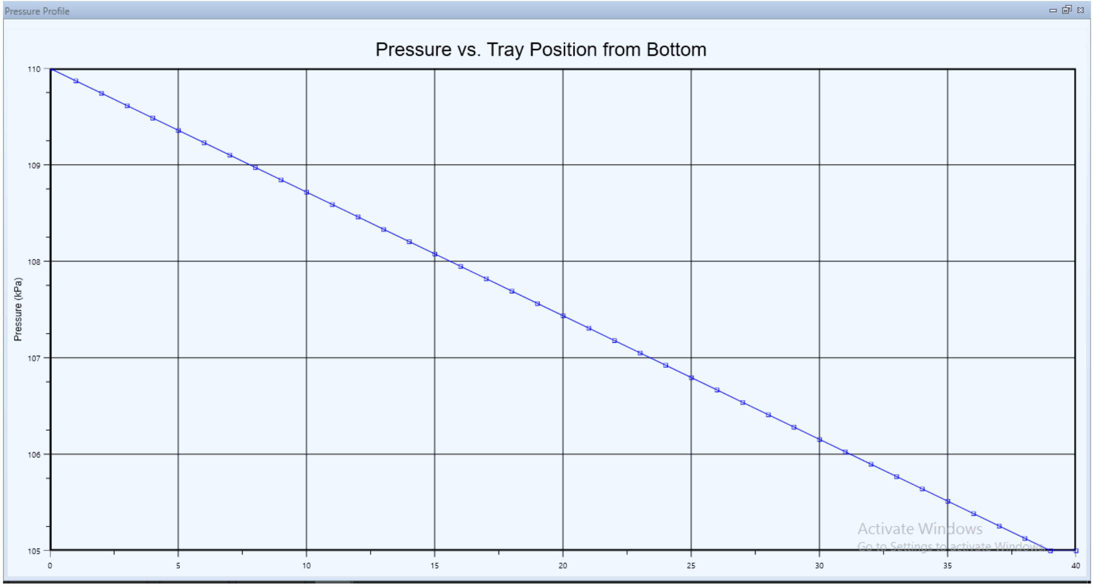

Figure (6.6) pressure profile

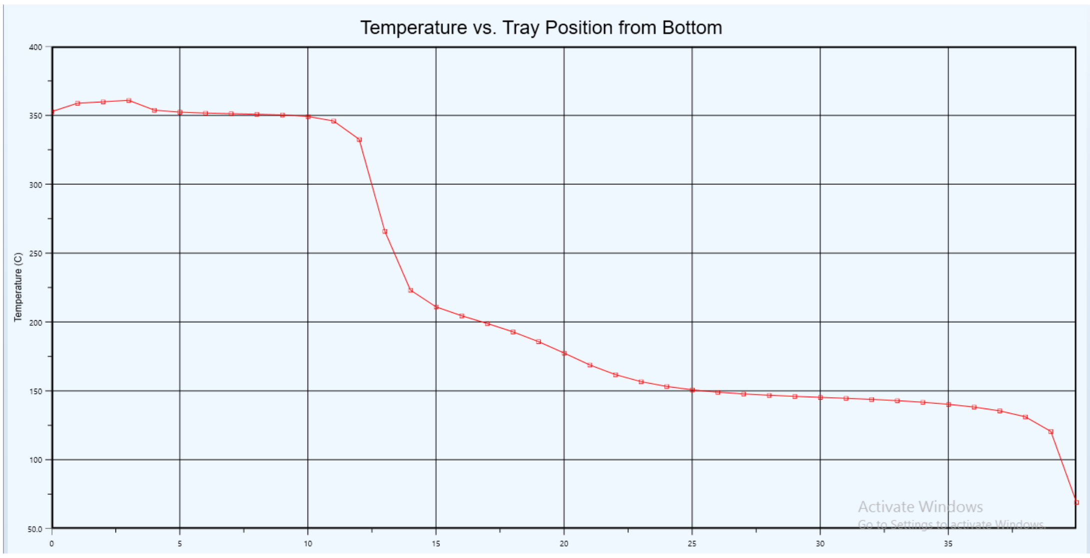

Figure (6.7) temperature profile

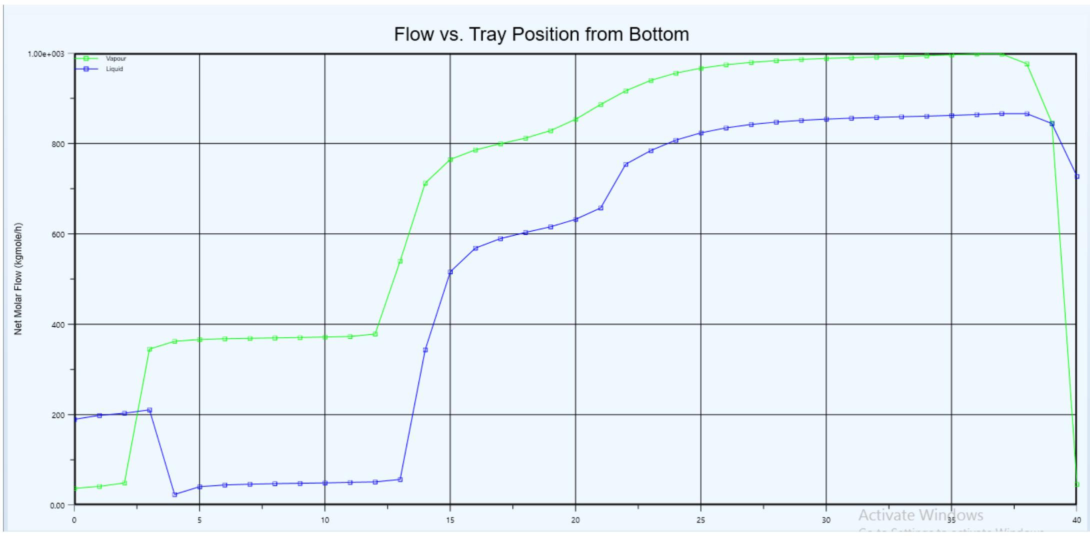

figure (6.8) Net flow profile

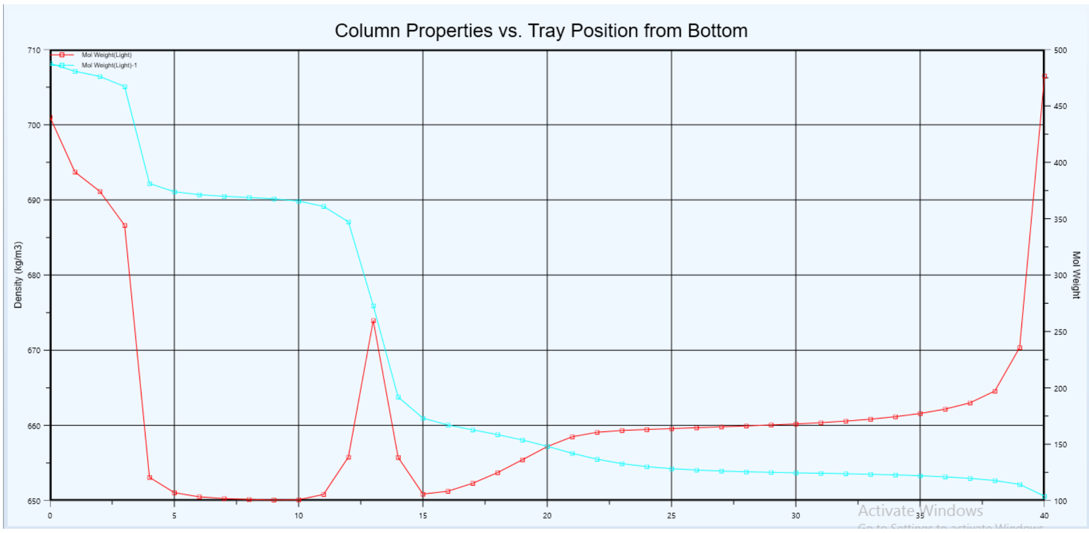

figure (6.9) column properties profile

6.3 comparison of production before and after expansion

Table (6.3) comparison of production before and after expansion

Product	Quantity before expansion Bbl/day	Quantity after expansion Bbl/day
Naphtha	900	24000
Kerosene	900	24000
Gasoil	3450	9200
Residue	9750	26000
Total	15000	40000

6.4. Cost estimation by Aspen HYHYS

6.4.1. Unite operation and equipment cost

Table (6.4) unite operation and equipment cost

Name	Equipment Cost [USD]	Installed Cost [USD]	Equipment Weight [LBS]	Total Installed Weight [LBS]
Preheater-2	66,200	216,200	14800	51769
Preheater-1	37,800	152,200	8700	34789
Atm Heater	199,400	562,100	54000	103646
Main Tower @T-100	485,800	1,199,700	153200	251199
Condenser @T-100-cond	30,200	131,100	7000	26633
Condenser @T-100-cond acc	27,100	153,600	5900	23281
Condenser @T-100-overhead s	0	0	0	0
Condenser @T-100-reflux pum	10,200	55,200	800	6637

6.4.2. Utilities cost

Table (6.5) Utilities cost

Name	Fluid	Rate	Rate Units	Cost per Hour	Cost Units
Electricity		95.7	KW	7.41675	USD/H
Cooling Water	Water	0.20312	MMGAL/H	24.3744	USD/H
Steam @100PSI	Steam	29.2557	KLB/H	238.141	USD/H

3

6.4.3. Cost summary

Table (6.6) cost summary

Total Capital Cost [USD]	5,383,390
Total Operating Cost [USD/Year]	3,575,920
Total Raw Materials Cost [USD/Year]	0
Total Product Sales [USD/Year]	0
Total Utilities Cost [USD/Year]	2,366,230
Desired Rate of Return [Percent/'Year]	20
P.O.Period [Year]	0
Equipment Cost [USD]	856,700
Total Installed Cost [USD]	2,470,100

Chapter Seven

Conclusion and recommendations

7.1. Conclusion

The simulation have been performed using specifications such as crude flow rate .the model is showed a lot of benefit and represented a useful results that have been used in the research .first of all material and energy balance have been calculated using some information from the model . All experiments are conducted using steady-state model developed under HYSYS environment and the results have been used first to design the distillation column of the crude distillation unit CDU of Obeid Refinery Company ORC to reach 41 trays ,diameter 1.5m and height 28.7m, where the results of naphtha is 6%, kerosene is 6%,gas oil 23% and residue 65% .In addition to that its used to design two heat exchangers with tube inside diameter 3\4 in , tube outside diameter 1 in and shell diameter 19 1\4 . Also the research used the program to estimate the cost of equipment and the operation cost of the units.

7.2. Recommendations

It is strongly recommend to add new crude distillation unit to satisfy shortage in domestic demand and achieve self-sufficiency.

After the expansion, it is strongly recommended to:

- Add catalytic reforming unit because the amount of naphtha has become economically sufficient to be treated locally rather than transfer it to Khartoum refinery.

- In future Add residue catalytic cracking unit because the amount of residue has become economically sufficient which is used as feed stock to produce valuable products.

Attachment

Table (1): Property of Nile Blend Crude

Properties	Result	Method
Density @ 15 ℃ , kg/m^3	877.1	ASTM D4052
Specific Gravity	0.8786	Calculation
API°	29.8	Calculation
Viscosity @100 ℃ , mm^2/s	8.856	ASTM D7024
Pour point , ℃	33.0	ASTM D5853
Carbon residue , m%	3.95	SH/T 0170-92
Water Content , m%	0.48	ASTM D4006
BS&W, m%	0.55	ASTM D4007
Salt content as NaCl, ppm	< 2.0	ASTM D6470
Acid number , mgKOH/g	0.88	ASTM D664
Sulfur , m%	0.057	ASTM D4294
Iron content mg/kg	5.8	ASTM D5708
Nickel mg/kg	9.6	
Calcium mg/kg	21.1	
Copper mg/kg	<0.1	
Sodium mg/kg	3.1	

Table (2): Property of Reforming Feed

Properties		Result	Method
Yield	m%	6.17	Calculation
	v %	7.41	Calculation
API°		62.3	Calculation
Density @ 15 ℃, kg/m^3		730.7	ASTM D4052
Specific Gravity		0.7302	Calculation
Acidity， mgKOH/g		0.04	ASTM D3242
Sulfur ,mg/kg		8.0	ASTM D5453
Nitrogen ,mg/kg		0.98	ASTM D4629
Copper ,µg/kg		<5.0	ASTM D3237
Lead ,µg/kg		<5.0	ASTM D3237
Arsenic ,µg/kg		<1.0	RIPP 65-90
Existent gum， mg/100mL		0.80	ASTM D381
Copper corrosion, grade		1a	ASTM D130
PNA， m%	P	61.46	SH/T 0239
	N	35.78	
	A	2.76	
Aromatic potential content， m%		22.61	Calculation
Distillation, ℃	IBP	54.1	ASTM D86
	10%	90.0	
	30%	106.6	
	50%	120.2	
	70%	133.6	ASTM D86
	90%	149.8	
	FBP	165.7	
Kuop		12.2	Calculation
BMCI		12.8	Calculation

Table(3): The detailed hydrocarbons of naphtha fraction

Carbon Number	Paraffins	Naphthenes	Aromatics	Total	Method
C3	0.02	-	-	0.02	SH/T 0239
C4	0.27	-	-	0.27	
C5	3.74	0.23	-	3.97	
C6	9.44	3.24	0.08	12.76	
C7	12.22	9.60	1.15	22.97	
C8	14.68	11.10	0.75	26.53	
C9	13.29	9.14	0.74	23.17	
C10	7.05	2.47	0.04	9.56	
>C11	0.75	-	-	0.75	
Total	61.46	35.78	2.76	100.0	
Benzene	0.08		Toluene	1.15	

Table (4): Property of kerosene fraction

Properties			Result	Method
Yield	m%		6.14	Calculation
	V%		6.75	Calculation
API°			45.6	Calculation
Density @ 15 ℃, kg/m^3			798.1	ASTM D4052
Specific Gravity			0.7988	Calculation
Viscosity mm^2/s		20 ℃	1.816	ASTM D7042
		40 ℃	1.320	
		-20 ℃	4.641	ASTM D445
Total Acidity ,mgKOH/g			0.034	ASTM D3242
Flash point(closed cup) ℃			58.0	ASTM D3828
Freezing point, ℃			-46.8	ASTM D5972
Aniline point , ℃			66.5	ASTM D611
Smoke point ,mm			20.0	ASTM D1322
Sulfur, mg/kg			30.0	ASTM D5453
Nitrogen , mg/kg			1.73	ASTM D4629
Mercaptan sulfur, mass%			N/A	ASTM D3227
Existent gum, mg/100mL			1.0	ASTM D381
Copper corrosion, grade			1a	ASTM D130
Hydrocarbon	Saturates %		92.2	ASTM D1319
	Olefin %		0.9	
	Aromatics %		6.9	

Table (5): Property of diesel fraction

Properties		Result	Method
Yield	m%	14.53	Calculation
	v%	15.17	Calculation
API°		36.73	Calculation
Density @ 15 ℃, kg/m^3		840.3	ASTM D4052
Specific Gravity		0.8404	Calculation
Viscosity , mm^2/s	20 ℃	6.661	ASTM D7042
	40 ℃	3.951	
Acidity , mgKOH/g		0.48	ASTM D3242
Flash point(closed cup) , ℃		126.0	ASTM D93
Cloud point , ℃		-1.1	ASTM D5773
Aniline point , ℃		79.8	ASTM D611
Sulfur, mg/kg		242.0	ASTM D5453

Table (6): Property of long residue, vacuum dist. & short residue

Properties		>350 ℃	>500 ℃	350-500 ℃	Method
Yield	m%	73.16	53.62	19.54	Calculation
	v%	70.77	50.99	19.73	Calculation
API°		24.29	21.90	31.11	Calculation
Density @ 15 ℃, kg/m^3		906.7	922.4	868.7	ASTM D4052
Specific Gravity		0.9083	0.9239	0.8702	Calculation
Viscosity @100 ℃,mm^2/s		31.730	66.270	4.495	ASTM D7042
Micro Carbon residue , m%		5.93	8.13	0.02	ASTM D4530
Pour point , ℃		39.0	40.0	43.0	ASTM D5853
Sulfur content , m%		1.09	0.092	0.055	ASTM D4294
Acid number mgKOH/g		1.03	0.95	1.27	ASTM D644
Ca, ppm		19.4	24.2	2.4	ASTM D5708
Fe, ppm		8.8	10.8	2.0	
Na, ppm		4.1	5.1	0.6	
V, ppm		0.6	0.7	0.1	
Ni, ppm		13.4	18.3	1.0	
Cu, ppm		<0.1	<0.1	<0.1	

Appendix

HYSYS report

Refluxed Absorber: T-100 @Main (continued)

CONDITIONS

Name		Atm Steam @Main	e oil to Atm twr @Main	Atm Resid @Main	flare gas @Main	Naphtha @Main
Vapour		0.0000	0.4728	0.0000	0.9999	0.0000
Temperature	(C)	176.6667 *	370.0000 *	352.7403	69.1186	69.0865
Pressure	(kPa)	1135.5389 *	273.6940 *	110.0000	105.0000	105.0000
Molar Flow	(kgmole/h)	27.7545	483.6297	189.5712	45.6804	72.6649
Mass Flow	(kg/h)	500.0000 *	145738.6031	92477.8813	2601.5103	7548.3234
Std Ideal Liq Vol Flow	(m3/h)	0.5010	165.6120	99.6689	3.7194	10.3286
Molar Enthalpy	(kJ/kgmole)	-2.741e+005	-3.619e+005	-6.392e+005	-1.603e+005	-2.174e+005
Molar Entropy	(kJ/kgmole-C)	86.45	1141	1816	166.5	134.3
Heat Flow	(kJ/h)	-7.6071e+06	-1.7500e+08	-1.2118e+08	-7.3230e+06	-1.5801e+07
Name		Sour Water @Main	Kerosene @Main	Diesel @Main	tm Condenser @Main	
Vapour		0.0000	0.0000	0.0000	---	
Temperature	(C)	69.0859	168.8225	265.8996	---	
Pressure	(kPa)	105.0000	107.3077	108.3333	---	
Molar Flow	(kgmole/h)	13.7874	64.1398	125.5157	---	
Mass Flow	(kg/h)	248.3812	9107.8312	34254.2292	---	
Std Ideal Liq Vol Flow	(m3/h)	0.2489	11.8550	40.2918	---	
Molar Enthalpy	(kJ/kgmole)	-2.828e+005	-2.610e+005	-4.253e+005	---	
Molar Entropy	(kJ/kgmole-C)	64.44	304.3	840.2	---	
Heat Flow	(kJ/h)	-3.8989e+06	-1.6743e+07	-5.3385e+07	3.5729e+07	

PROPERTIES

Name		Atm Steam @Main	Crude oil to Atm twr @M	Atm Resid @Main	flare gas @Main	Naphtha @Main
Molecular Weight		18.02	301.3	487.8	56.95	103.9
Molar Density	(kgmole/m3)	48.75	0.1100	1.437	3.768e-002	6.801
Mass Density	(kg/m3)	878.2	33.16	701.0	2.146	706.5
Act. Volume Flow	(m3/h)	0.5694	4395	131.9	1212	10.68
Mass Enthalpy	(kJ/kg)	-1.521e+004	-1201	-1310	-2815	-2093
Mass Entropy	(kJ/kg-C)	4.799	3.787	3.724	2.923	1.293
Heat Capacity	(kJ/kgmole-C)	84.97	936.5	1510	101.8	229.4
Mass Heat Capacity	(kJ/kg-C)	4.717	3.108	3.096	1.787	2.208
LHV Molar Basis (Std)	(kJ/kgmole)	0.0000	---	---	---	---
HHV Molar Basis (Std)	(kJ/kgmole)	4.101e+004	---	---	---	---
HHV Mass Basis (Std)	(kJ/kg)	2276	---	---	---	---
CO2 Loading		---	---	---	---	---
CO2 Apparent Mole Conc.	(kgmole/m3)	---	---	---	---	---
CO2 Apparent Wt. Conc.	(kgmol/kg)	---	---	---	---	---
LHV Mass Basis (Std)	(kJ/kg)	0.0000	---	---	---	---
Phase Fraction [Vol. Basis]		0.0000	0.2800	0.0000	1.000	0.0000
Phase Fraction [Mass Basis]		0.0000	0.2525	0.0000	1.000	0.0000
Phase Fraction [Act. Vol. Basis]		0.0000	0.9626	0.0000	1.000	0.0000
Mass Exergy	(kJ/kg)	131.7	338.6	284.2	43.95	6.404
Partial Pressure of CO2	(kPa)	0.0000	0.0000	0.0000	0.0000	0.0000
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000	0.0000
Act. Gas Flow	(ACT_m3/h)	---	4230	---	1212	---
Avg. Liq. Density	(kgmole/m3)	55.40	2.920	1.902	12.28	7.035
Specific Heat	(kJ/kgmole-C)	84.97	936.5	1510	101.8	229.4
Std. Gas Flow	(STD_m3/h)	656.2	1.144e+004	4482	1080	1718
Std. Ideal Liq. Mass Density	(kg/m3)	998.0	880.0	927.9	699.4	730.8
Act. Liq. Flow	(m3/s)	1.582e-004	4.565e-002	3.664e-002	1.835e-008	2.968e-003
Z Factor		6.229e-003	---	1.471e-002	---	---
Watson K		---	11.85	12.05	12.16	12.07
User Property		---	---	---	---	---
Partial Pressure of H2S	(kPa)	0.0000	0.0000	0.0000	0.0000	0.0000
Cp/(Cp - R)		1.108	1.009	1.006	1.089	1.038
Cp/Cv		1.198	1.006	1.659	1.096	1.179
Ideal Gas Cp/Cv		1.314	1.010	1.006	1.090	1.048
Ideal Gas Cp	(kJ/kgmole-C)	34.79	870.0	1377	101.1	182.4
Mass Ideal Gas Cp	(kJ/kg-C)	1.931	2.887	2.824	1.775	1.756
Heat of Vap.	(kJ/kgmole)	3.638e+004	5.422e+005	6.268e+005	4.772e+004	4.479e+004

Refluxed Absorber: T-100 @Main (continued)

PROPERTIES

Name		Atm Steam @Main	Crude oil to Atm twr @M	Atm Resid @Main	flare gas @Main	Naphtha @Main
Kinematic Viscosity	(cSt)	0.1734	---	0.5483	---	0.4760
Liq. Mass Density (Std. Cond)	(kg/m3)	1015	880.0	943.1	741.5	753.3
Liq. Vol. Flow (Std. Cond)	(m3/h)	0.4927	165.6	98.05	3.508	10.02
Liquid Fraction		1.000	0.5272	1.000	7.814e-005	1.000
Molar Volume	(m3/kgmole)	2.051e-002	9.087	0.6959	26.54	0.1470
Mass Heat of Vap.	(kJ/kg)	2019	1799	1285	837.8	431.2
Phase Fraction [Molar Basis]		0.0000	0.4728	0.0000	0.9999	0.0000
Surface Tension	(dyne/cm)	43.00	9.678	12.26	64.32	16.61
Thermal Conductivity	(W/m-K)	0.6782	---	9.053e-002	---	0.1052
Bubble Point Pressure	(kPa)	925.2	1657	110.0	911.6	104.9
Viscosity	(cP)	0.1523	---	0.3844	---	0.3363
Cv (Semi-Ideal)	(kJ/kgmole-C)	76.65	928.2	1502	93.46	221.1
Mass Cv (Semi-Ideal)	(kJ/kg-C)	4.255	3.080	3.079	1.641	2.128
Cv	(kJ/kgmole-C)	70.94	931.1	910.7	92.85	194.6
Mass Cv	(kJ/kg-C)	3.938	3.090	1.867	1.630	1.873
Cv (Ent. Method)	(kJ/kgmole-C)	---	---	1097	---	---
Mass Cv (Ent. Method)	(kJ/kg-C)	---	---	2.248	19.26	10.56
Cp/Cv (Ent. Method)		7.747e-002	0.8539	1.377	9.280e-002	0.2092
Reid VP at 37.8 C	(kPa)	---	20.85	6.387e-004	279.9	28.74
True VP at 37.8 C	(kPa)	---	57.30	6.468	618.7	37.59
Liq. Vol. Flow - Sum(Std. Cond)	(m3/h)	0.4927	163.5	98.05	3.508	10.02
Viscosity Index		---	-27.67	-6.580	-14.39	-9.545

Name		Sour Water @Main	Kerosene @Main	Diesel @Main		
Molecular Weight		18.02	142.0	272.9		
Molar Density	(kgmole/m3)	54.02	4.637	2.470		
Mass Density	(kg/m3)	973.2	658.5	674.0		
Act. Volume Flow	(m3/h)	0.2552	13.83	50.83		
Mass Enthalpy	(kJ/kg)	-1.570e+004	-1838	-1559		
Mass Entropy	(kJ/kg-C)	3.577	2.143	3.079		
Heat Capacity	(kJ/kgmole-C)	78.19	372.0	783.7		
Mass Heat Capacity	(kJ/kg-C)	4.340	2.620	2.872		
LHV Molar Basis (Std)	(kJ/kgmole)	7.217e-004	---	---		
HHV Molar Basis (Std)	(kJ/kgmole)	4.101e+004	---	---		
HHV Mass Basis (Std)	(kJ/kg)	2276	---	---		
CO2 Loading		---	---	---		
CO2 Apparent Mole Conc.	(kgmole/m3)	---	---	---		
CO2 Apparent Wt. Conc.	(kgmol/kg)	---	---	---		
LHV Mass Basis (Std)	(kJ/kg)	4.006e-005	---	---		
Phase Fraction [Vol. Basis]		0.0000	0.0000	0.0000		
Phase Fraction [Mass Basis]		0.0000	0.0000	0.0000		
Phase Fraction [Act. Vol. Basis]		0.0000	0.0000	0.0000		
Mass Exergy	(kJ/kg)	12.85	62.81	162.0		
Partial Pressure of CO2	(kPa)	0.0000	0.0000	0.0000		
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000		
Act. Gas Flow	(ACT_m3/h)	---	---	---		
Avg. Liq. Density	(kgmole/m3)	55.40	5.410	3.115		
Specific Heat	(kJ/kgmole-C)	78.19	372.0	783.7		
Std. Gas Flow	(STD_m3/h)	326.0	1517	2968		
Std. Ideal Liq. Mass Density	(kg/m3)	998.0	768.3	850.2		
Act. Liq. Flow	(m3/s)	7.089e-005	3.842e-003	1.412e-002		
Z Factor		6.831e-004	6.297e-003	9.788e-003		
Watson K		19.47	12.10	12.05		
User Property		---	---	---		
Partial Pressure of H2S	(kPa)	0.0000	0.0000	0.0000		
Cp/(Cp - R)		1.119	1.023	1.011		
Cp/Cv		1.167	1.176	1.011		
Ideal Gas Cp/Cv		1.326	1.028	1.012		
Ideal Gas Cp	(kJ/kgmole-C)	33.85	310.5	694.2		
Mass Ideal Gas Cp	(kJ/kg-C)	1.879	2.187	2.544		

* Specified by user.

Company Name Not Available
Bedford, MA
USA

Case Name: Obied Refinery.hsc

Unit Set: SI

Date/Time: Wed Feb 16 10:54:53 2022

Refluxed Absorber: T-100 @Main (continued)

PROPERTIES

Name		Sour Water @Main	Kerosene @Main	Diesel @Main		
Heat of Vap.	(kJ/kgmole)	4.096e+004	4.393e+004	1.348e+005		
Kinematic Viscosity	(cSt)	0.4167	0.3483	0.2926		
Liq. Mass Density (Std. Cond)	(kg/m3)	1015	786.8	867.7		
Liq. Vol. Flow (Std. Cond)	(m3/h)	0.2448	11.58	39.48		
Liquid Fraction		1.000	1.000	1.000		
Molar Volume	(m3/kgmole)	1.851e-002	0.2156	0.4049		
Mass Heat of Vap.	(kJ/kg)	2274	309.4	494.1		
Phase Fraction [Molar Basis]		0.0000	0.0000	0.0000		
Surface Tension	(dyne/cm)	64.33	11.92	11.49		
Thermal Conductivity	(W/m-K)	0.6616	9.849e-002	9.421e-002		
Bubble Point Pressure	(kPa)	29.91	107.3	108.3		
Viscosity	(cP)	0.4055	0.2293	0.1972		
Cv (Semi-Ideal)	(kJ/kgmole-C)	69.88	363.7	775.4		
Mass Cv (Semi-Ideal)	(kJ/kg-C)	3.879	2.561	2.841		
Cv	(kJ/kgmole-C)	66.98	316.3	775.4		
Mass Cv	(kJ/kg-C)	3.718	2.227	2.841		
Cv (Ent. Method)	(kJ/kgmole-C)	66.70	312.8	---		
Mass Cv (Ent. Method)	(kJ/kg-C)	60.88	7.724	4.019		
Cp/Cv (Ent. Method)		7.130e-002	0.3392	0.7146		
Reid VP at 37.8 C	(kPa)	---	1.231	0.4077		
True VP at 37.8 C	(kPa)	---	2.894	2.688		
Liq. Vol. Flow - Sum(Std. Cond)	(m3/h)	0.2448	11.58	39.48		
Viscosity Index		-12.70	-17.78	-24.26		

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