ا لاستـهـلال

قال تعالى:

قَالُواْ سُبْحَانَكَ لاَ عِلْمَ لَنَا إِلاَّ مَا عَلَّمْتَنَا إِنَّكَ أَنتَ الْعَلِيمُ الْحَكِيمُ(32))

سورة البقره

DEDICATION

To our parents and to our families who made this accomplishment possible

ACKNOWLEDGMENTS

First and foremost, we would like to thank the chairman of our committee, Mr. Abdelgadir Bashir Banaga, for his support, outstanding guidance and encouragement throughout our senior project.

We would like to express our gratitude and appreciation to Dr. Sumia Abd Almoniam for all the help and guidance he provided throughout our education.

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Finally we would like to thank our families, especially our parents, for their encouragement, patience, and assistance over the years. We are forever indebted to our parents, who have always kept us in their prayers

Abstract

This project present anew methodology to design the distillation column based on simulation of the unit using ASPEN HYSYS, the methodology consider that the Nile blend is a complex mixture so it take every two products and from the simulation results determent the light and heavy key between the two product to calculate the minimum number of try and the minimum reflux to determine the actual number of try and the dimensions of the column in general , the results of the design is 52 trays and diameter of 2.191 m and 26.6m high . Also the project concerned on control of the unit using different type of controller. using the simulation model to control the production of the products and try to increasing the productivity of naphtha from 7.75% to 11.62% ,and increasing the kerosene from 3.87% to 5.8% .experiment of gas chromatography is been done in CPL to use it in validation of model.

Key word: distillation, simulation, design, control, experiment

المستخلص

المشروع يوضح منهجيه جديدة لتصميم عمود التقطير الجوى استنادا على النتائج المتحصل عليها من عملية المحاكاة التى تم اجراءها على وحدة تقطير الخام في مصفاة الخرطوم باستخدام برنامج الهايسس , وهذه الطريقه تقوم بأخذ كل منتجين لتحديد المفاتيح الخفيفة والثقيلة بينهما من نتائج المحاكاة لتحسب منها الحد الادنى لعدد الصوانى و الراجع ومنها تقوم بحساب عدد الصوانى الكلية و ابعاد العمود وقد تم الحصول على نتائج عدد 52 صينيه و قطر 2متر وارتفاع 26.6 متر . وقد تم اجراء عدد من عمليات التحكم على النموذج باستخدام البرنامج للتحكم في انتاجية بعض منتجات الوحدة لزيادة انتاجية النافثا من 7.75% الى 11.62% وزيادة نسبة انتاجية الكيروسين من 3.87% الى 5.8% . ايضاً تم اجراء تجارب كروماتوغرافيا الغاز على بعض المنتجات في مختبرات النفط المركزيه وذلك للمقارنه بينها والنموذج المصصم للتاكد من صحته .

Table Of Contents

1. INTRODUCTION

2.LITERATURE REVIEW

3. METHODOLOGY

4. MATERIAL BALANCE AND ENERGY BALANCE

5.DESIGN OF CRUDE DISTILLATION TOWER

6.CONTROL SYSTEM

7. COST ESTIMATION

8. RESULTS AND DISCUSSION

LIST OF TABLES

LIST OF FIGURES

Chapter 1

Introduction

1.1. Background

1.1.2. 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. 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 .

1.2. Project objective

1 Simulation of the crude distillation unit CDU of Khartoum refinery company KRC using Aspen HYSYS program.
2. Design of distillation column using result from the simulation model.
3. Control the unit using a different type of controller in the software.
4. Increasing the production of some products, naphtha and kerosene.

1.3. Methodology summary

We will build the (KRC) 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.

Chapter 2

Literature review

2.1 Previous work

In 2004, 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 (Nuhu M and etal, 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 ASPEN^{TM} HYSYS 7.1. (A.Rahman and K. Kirtania,2011) a retrofit design methodology and simulation framework for heat-integrated crude oil distillation systems was studies by using HYSYS (Mamdouh Gadallaa 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 (E. O. Okeke& A. A. Osakwe-Akofe , 2009). 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* (SampathYela ,2009).

2.2. Crude Distillation Unit

2.2.1. 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 desalter 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 desalter 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 unvaporized portion of the crude leaves the bottom of the tower via a steam stripper section, 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. Distillate products are removed from selected trays (draw-off trays) in this sections of the tower. These streams are stream stripped and sent to storage. 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 this stream is returned as reflux while the remainder is delivered to the light end processes for stabilizing and further distillation.

The side stream distillates shown in the diagram are:

1. Heavy gas oil (has the highest Boiling Point)
2. Light gas oil (will become Diesel)
3. Kerosene (will become Jet Fuel)

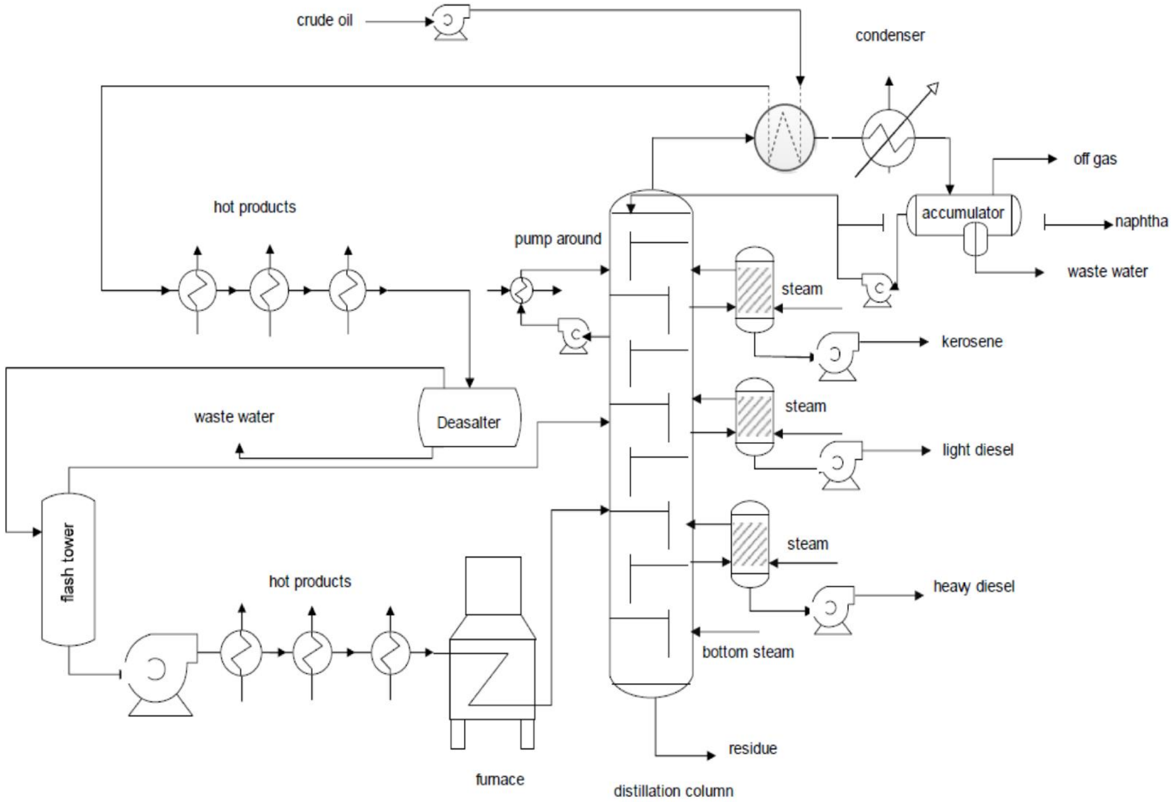


Figure 2.1: Atmospheric distillation unit

A 'Pump around' section is included at the light gas oil draw off. This is simply an internal condenser which takes heat out of that section of the tower. This in turn ensures a continued reflux stream flow below that section. The product side streams are stripped free of entrained light ends in separate stripping towers. These towers also contain fractionation trays (usually four but sometimes as many as six) and the side stream drawn off the main tower enters the top tray of its respective stripper. steam is injected below the bottom tray and moves up the tower to leave at the top, together with the light ends strip out, and is returned to the main fractionators at a point directly above the side stream draw-off tray. These side stream stripper towers are usually stacked one above the other in a single column in such a way as to allow free flow from the side stream draw-off tray to its stripper tower. on a few occasions, where the particular side stream specification requires it, the stripping may be effected by reboiling instead of using steam one such requirement maybe in the kero side stream if this stream is to be routed directly into jet fuel blending and therefore must be dry. The residue (unvaporized portion of the crude) leaves the flash zone to flow over four stripping trays counter current to the flow of stripping steam. This stripping steam enters the tower below the bottom stripping tray. Its purpose primarily is to strip the residue free of entrained light ends. The fact that this steam enters the flash zone it also enhances the flashing of the crude in this zone by creating a reduced partial pressure for the liquid/vapor separation. This becomes an important factor in the design and operation of the atmospheric crude distillation unit. The stripped residue leaves the bottom of the unit to be routed either through the unit's heat exchanger system and the to product storage or hot to some downstream processing unit such as a vacuum distillation unit or a thermal cracker. (Daved s 2006).

2.2.4. Products of Crude distillation unit

In the order of increasing boiling points, the main products from a typical crude distillation unit are as follows:-

Fuel gas. The fuel gas consists mainly of methane and ethane. In some refineries, propane in excess of LPG requirements is also included in the fuel gas stream. This stream is also referred to as ‘‘dry gas.’’ (C1, C2, C3)

Wet gas. The wet gas stream contains propane and butanes as well as methane and ethane. The propane and butanes are separated to be used for LPG and, in the case of butanes, for gasoline blending and alkylation unit feed. (C2, C3, C4)

LSR naphtha. The stabilized LSR naphtha (or LSR gasoline) stream is desulfurized and used in gasoline blending or processed in an isomerization unit to improve octane before blending into gasoline. (C5, C6)

HSR naphtha or HSR gasoline. The naphtha cuts are generally used as catalytic reformer feed to produce high-octane reformate for gasoline blending and aromatics. (C7 to C10)

Kerosene The kerosene stream is treated and then sent to the blending pool for sale as kerosene product. (C9 toC15)

Diesel The diesel cut is treated and then sent to the blending pool for sale as diesel fuel. (C13 to C18)

Gas oils. The light, atmospheric and vacuum gas oils arm processed in a hydrocracker or catalytic cracker to produce gasoline, jet, and diesel fuels. The heavier vacuum gas oils can also be used as feedstock for lubricating oil processing units. (C13 to C45)

Residue. The vacuum still bottoms can be processed in a visbreaker, Coker, or deasphalting unit to produce heavy fuel oil or cracking and/or lube base stocks. For asphalt crudes, the residuum can be processed further to produce road and/or roofing asphalts (C40 and up)

2.3. 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.

2.3.1. 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.

2.3.2. 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

2.3.3. Oil Characterization

The petroleum characterization method in HYSYS converts your 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 hypocomponent 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

The characterization procedure performs its calculations based on an internally calculated TBP curve. If you supply an ASTM or EFV distillation curve, it is converted to a TBP curve using standard methods described in the API Data Book. If you do not supply any distillation data, then an average TBP distillation curve is generated for you based on the overall molecular weight, density, and Watson (UOP) K factor of your fluid.(HYSYS manual,2004).

Chapter 3

Methodology

This chapter will descript the simulation steps for CDU or the procedure of the simulation by HYSYS, and also provide the procedure that's followed to design the distillation tower, the procedure of increasing some of the products.

3.1. Case study:-

The crude distillation unit of Sudan Khartoum refinery co., ltd. is mainly used to process the mixed crude oil of HEGLIG, unity, and TOMA etc. of Sudan, with its crude oil processing capability of 2.5 million tons per year, which produce the fuels and the intermediates of other unit as well as had gotten once through succeeded in February of 2000 after passing through a preheat train, 294.1m3 /h of Nile blend crude is fed into a pre-flash separator operating at 193°c and 75 psia. The flashed streams are then fed to the atmospheric crude column. Three coupled side strippers, and three pumps around circuits. A naphtha product is produced overhead, a kerosene product is produced from the first side stripper, a diesel product is produced from the second side stripper, and atmospheric gas oil (ago) is produced from the third side stripper. Both the ago side stripper and the diesel side stripper are 'steam stripped', while the kerosene side stripper has a reboiler .

3.2. process simulation procedure:

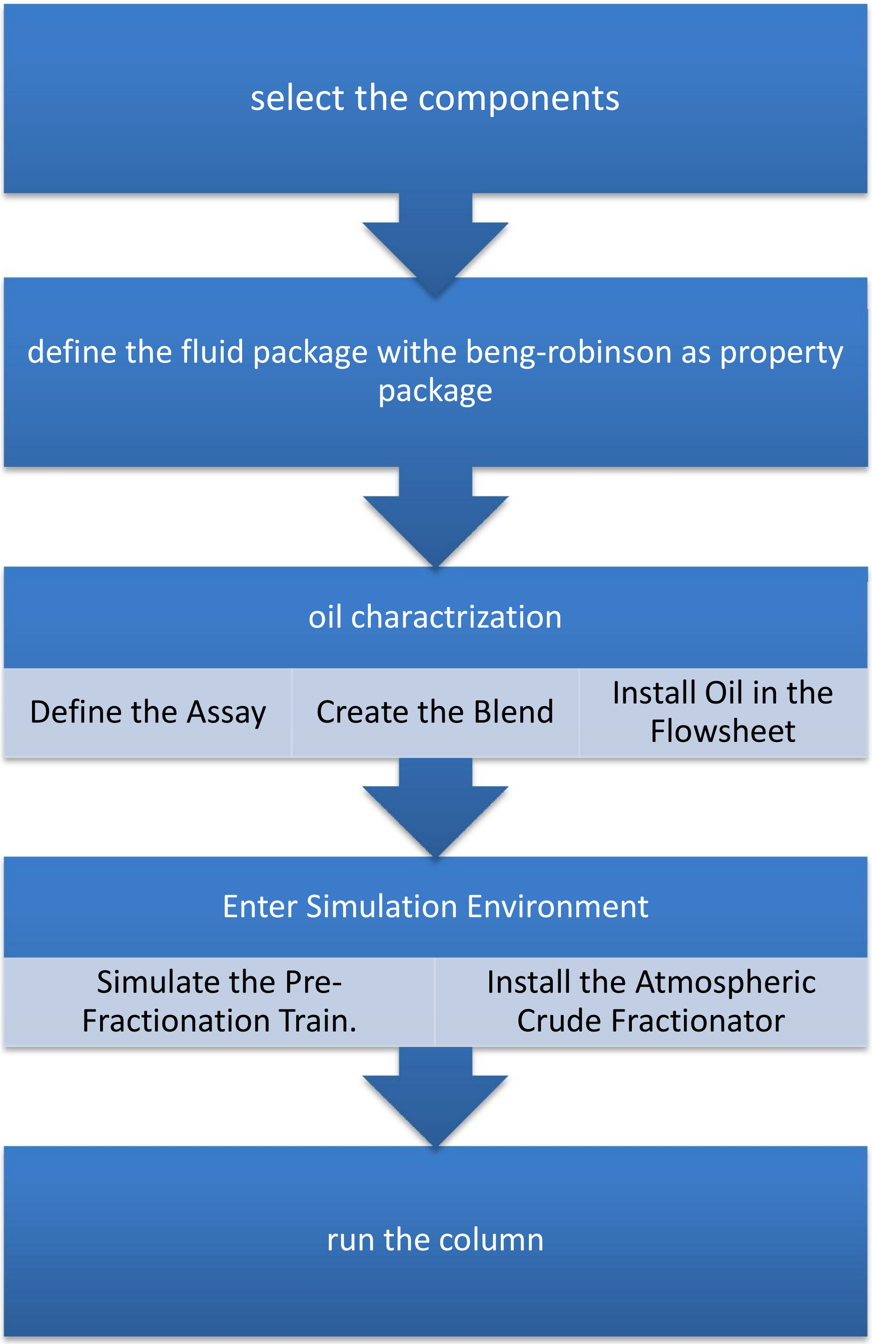


Figure 3.1: process simulation procedure

3.3. Multi component distillation

3.3.1. 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. (R.K.Sinnott,2005)

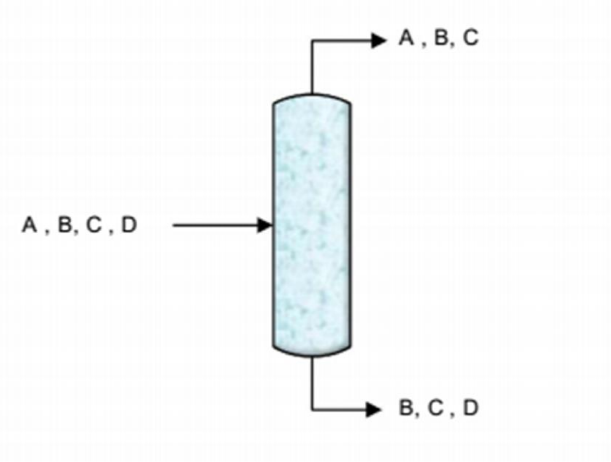


Figure 3.2: distribution of component between top and bottom

3.3.2. 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.

$$N_{min} = \frac{\log\left(\frac{X_{D,l}}{X_{B,H}} \times \frac{X_{B,H}}{X_{B,l}}\right)}{\log(\alpha_{l,H})} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3.1)$$

Where the mean relative volatility is approximated by

$$\alpha_{L,H} = \frac{\alpha_{,lk}}{\alpha_{,Hk}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots(3.2)$$

3.3.3. Minimum reflux ratio, using Underwood's method

For feed conditions where the relative volatilities remain constant, UNDERWOOD developed the following two equations from which R_{min}may be calculated:

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

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

- where: xfA ,xfB ,xfC ,xdA ,xdB ,xdC, 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, as in equation 3.3, αA,αB,αC, etc., are the volatilities with respect to the least volatile component,
- θ is the root of equation 3.3, which lies between the values of αA and αB. 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. (R.K.Sinnott,2005)

3.3.4. Number of plates at total reflux

Capital, and interest-on-capital costs, are related to the number of stages, whereas operating costs are tied to reflux ratio and thus to fuel costs for providing heat to the reboiler. Less reflux is needed if more stages are added, so operating costs decrease as capital costs increase. For a specified separation, the reflux ratio and equilibrium stages must be greater than their minimum values. The actual reflux ratio should be established by economic considerations at some multiple of minimum reflux. The corresponding number of stages is then determined by suitable analytical or graphical methods.(D.J.Seader, 2011)

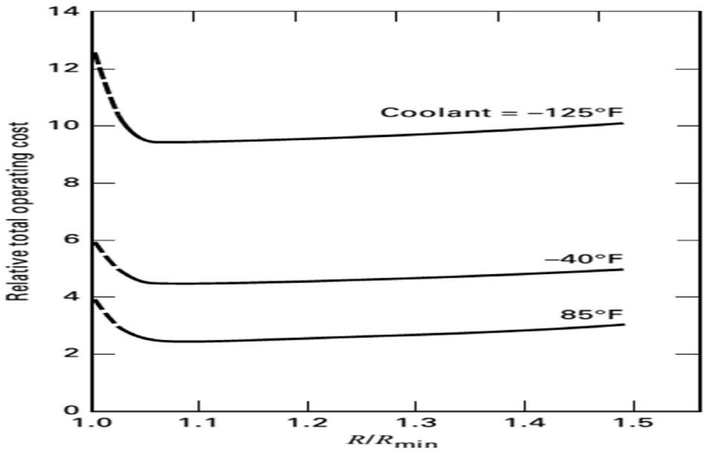


Figure 3.3: Effect of reflux ratio on cost.

3.3.5. 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 R_{min} and the number of plates at total reflux N_{min} are required. This is shown in Figure3.4, where the group $[(N+1) - (N_{min}+1)]/(N+2)$is plotted against $(R-R_{min})/(R+1)$. (Richardson and J.H.Harker,2002).

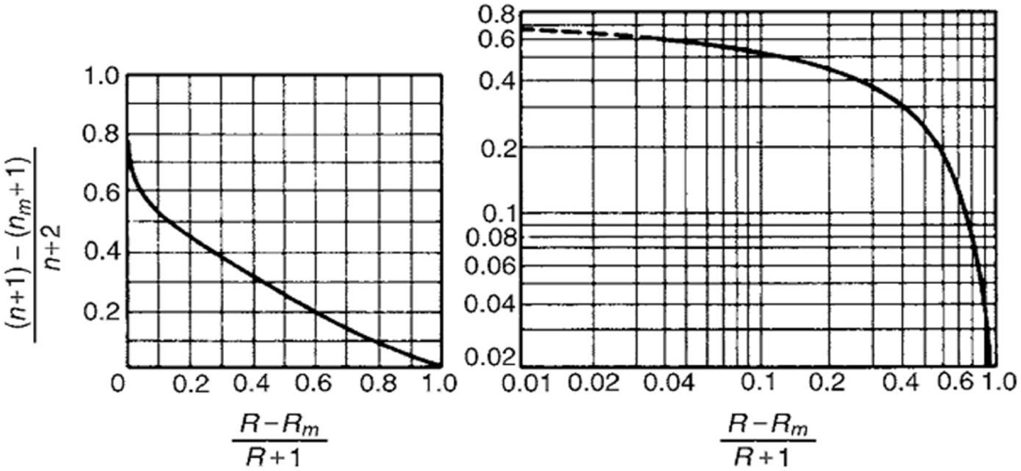


Figure 3.4: Relation between reflux ratio and number of plates.

3.4. 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. (R.K.Sinnott,2005).

3.4.1. 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. (R.K.Sinnott,2005)

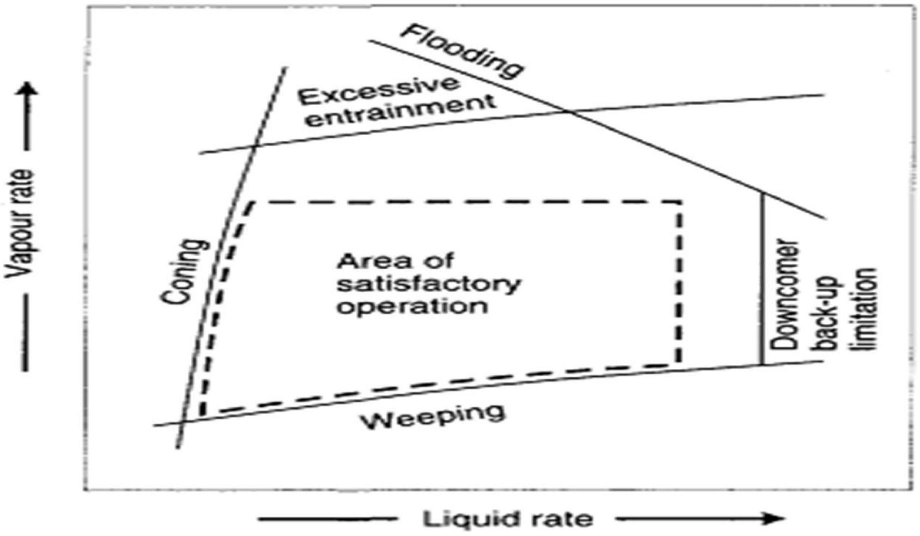


Figure 3.5: Sieve plate performance diagram

3.4.2. 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: -

$$\textbf{Efficiency} = 0.17 - 0.616 \log_{10}(\mu) \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots..(3.5)$$

Where

μL is the viscosity at the mean tower temperature. (R.K.Sinnott,2005).

3.4.3. Determination of the Column Diameter

Flow Parameter

$$F_{LV} = \left(\frac{L_n}{V_n}\right)\left(\frac{\rho_v}{\rho_L}\right)^{0.5} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots..(3.6)$$

F_{LV} = Liquid Vapor Factor

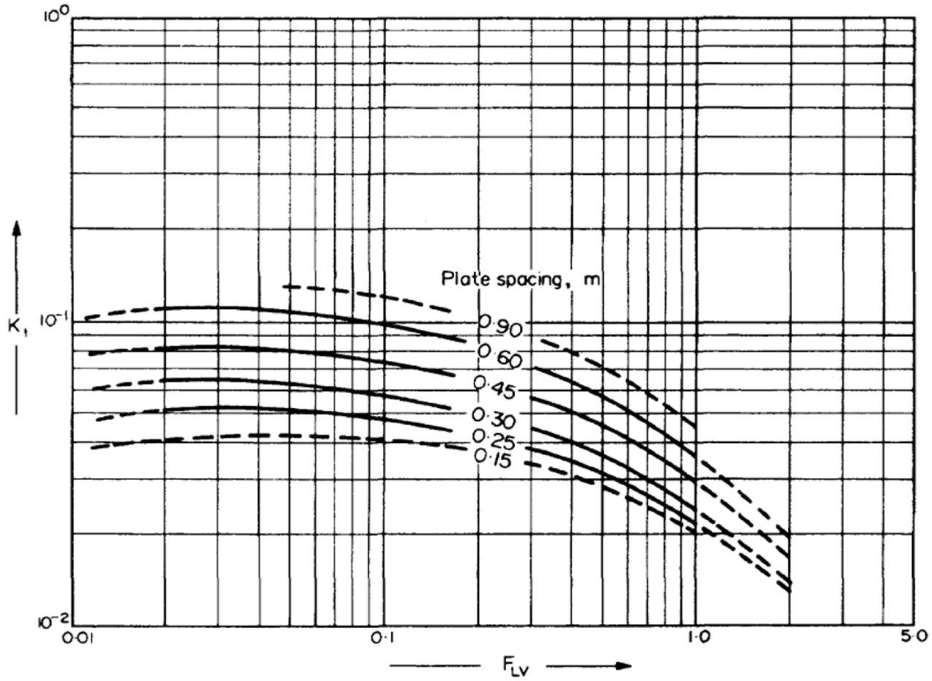


Figure 3.6: Flooding velocity, sieve plates

Capacity Parameter

$$V_{nf} = C_{sb}\left(\frac{\sigma}{20}\right)^{0.2}\left(\frac{\rho l - \rho v}{\rho v}\right)^{0.5} \text{.........................(3.7)}$$

Net column area used in separation is $A = Q_v/V_n$...............(3.8)

Volumetric flow rate of vapors = Q_v

Now, net area An= Qv/V_n

Active area Ac = $A_n + A_d$

Where, A_d = down comer area.

So Diameter of Column Is

$$A_c=(\pi/4)D^2$$

$$D = \sqrt{(4Ac/\pi)}\ .$$

3.5. Petroleum and complex mixture distillation

Although the principles of multicomponent distillation apply to petroleum, synthetic crude oil, and other complex mixtures, this subject warrants special consideration for the following reasons:

1- Such feedstock's are of exceedingly complex composition, consisting of, in the case of petroleum, many different types of hydrocarbons and perhaps of inorganic and other organic compounds. The number of carbon atoms in the components may range from 1 to more than 50, so that the compounds may exhibit atmospheric-pressure boiling points from −162°C (−259°F) to more than 538°C (1000°F). In a given boiling range, the number of different compounds that exhibit only small differences in volatility multiplies rapidly with increasing boiling point. For example, 16 of the 18 octane isomers boil within a range of only 12°C (22°F).

2- Products from the distillation of complex mixtures are in themselves complex mixtures. The character and yields of these products vary widely, depending upon the source of the feedstock. Even crude oils from the same locality may exhibit marked variations.

3- The scale of petroleum-distillation operations is generally large, and as discussed in detail by Nelson (Petroleum Refinery Engineering, 4th ed., McGraw-Hill, New York, 1958) and Watkins (Petroleum Refinery Distillation, 2d ed., Gulf, Houston, 1979), such operations are common in several petroleum refinery processes including atmospheric distillation of crude oil, vacuum distillation of bottoms residuum obtained from atmospheric distillation, main fractionation of gaseous effluent from catalytic cracking of various petroleum fractions, and main fractionation of effluent from thermal coking of various petroleum fractions. These distillation operations are conducted in large pieces of equipment that can consume large quantities of energy. Therefore, optimization of design and operation is very important and frequently leads to a relatively complex equipment configuration (M.F.DOHTRY.et al,2008) .

3.5.1. Design procedure

Two general procedures are available for designing fractionators that Process petroleum, synthetic crude oils, and complex mixtures. The first, which was originally developed for crude units by Packie [Trans. Am. Inst.Chem. Eng. J., 37, 51 (1941)], extended to main fractionators by Houghland, Lemieux, and Schreiner [Proc. API, sec. III, Refining, 385 (1954)], and further elaborated and described in great detail by Watkins (op. cit.), uses material and energy balances, with empirical correlations to establish tray requirements, and is essentially a hand calculation procedure that is a valuable learning experience and is suitable for preliminary designs. Also, when backed by sufficient experience from previous designs, this procedure is adequate for final design. In the second procedure, which is best applied with a digital computer, the complex mixture being distilled is represented by actual components at the light end and by perhaps 30 pseudo components (e.g., petroleum fractions) over the remaining portion of the TBP distillation curve for the column feed. Each of the pseudo components is characterized by a TBP range, an average normal boiling point, an average API gravity, and an average molecular weight. Rigorous material balance, energy balance, and phase equilibrium calculations are then made by an appropriate equation-tearing method, as shown by Cecchetti et al. [Hydrocarbon Process., 42(9), 159 (1963)] or a simultaneous-correction procedures shown, e.g., by Goldstein and Stanfield [Ind. Eng. Chem. Process Des. Dev., 9, 78 (1970)] and Hess et al. [Hydrocarbon Process, 56(5), 241 (1977)]. Highly developed procedures of the latter type, suitable for preliminary or final design, are included in most computer-aided steady-state process design and simulation programs as a special case of interlinked distillation, wherein the crude tower or fractionator is converged simultaneously with the side cut stripper columns.

Regardless of the procedure used, certain initial steps must be taken for the determination or specification of certain product properties and yields based on the TBP distillation curve of the column feed, method of providing column reflux, column-operating pressure type of condenser, and type of side cut strippers and stripping requirements. (M.F.DOHTRY.et al,2008)

3.6. Methodology of increasing naphtha and kerosene production

Naphtha
On the monitor we make these changes

Table 3.1: Simulation specification in the HYSYS monitor (naphtha)

Specifications	Active	Estimate
Reflux Ratio	✓	✓
Distillate Rate	✓	✓
Reflux Rate		✓
Vap Prod Rate	✓	✓
Btms Prod Rate		✓
kero Prod Flow	✓	✓
kero Boil Up Ratio	✓	✓
light diesel Prod Flow	✓	✓
PA_1_Rate(Pa)	✓	✓
PA_1_Dt(Pa)	✓	✓
PA_2_Rate(Pa)		✓
PA_2_Dt(Pa)	✓	✓
PA_3_Rate(Pa)	✓	✓
PA_3_Dt(Pa)	✓	✓
heavy diesel ss Prod Flow	✓	✓

Kerosene

Table 3.2: Simulation specification in the HYSYS monitor (kerosene)

Specifications	Active	Estimate
Reflux Ratio	✓	✓
Distillate Rate	✓	✓
Reflux Rate		✓
Vap Prod Rate	✓	✓
Btms Prod Rate		✓
kero Prod Flow	✓	✓
kero Boil Up Ratio	✓	✓
light diesel Prod Flow	✓	✓
PA_1_Rate(Pa)	✓	✓
PA_1_Dt(Pa)	✓	✓
PA_2_Rate(Pa)		✓
PA_2_Dt(Pa)	✓	✓
PA_3_Rate(Pa)	✓	✓
PA_3_Dt(Pa)	✓	✓
heavy diesel ss Prod Flow	✓	✓

Chapter 4

Material balance and energy balance

4.1. Material balance

4.1.1. Mass balance around Flash tower

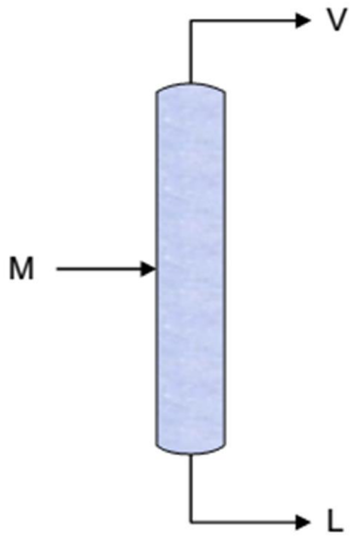


Figure 4.1: flash tower

Total flow =258 t/hr

$$\frac{V}{M} = 0.0027$$

<u>At top</u>

V = 258 * 0.0027 =0.6944 t/hr

<u>At bottom</u>

B = 0.9973 * 258 = 257.3 T/hr

4.1.2. Material balance around distillation column

The equation of material balance for any system:-

Input +generation –consumption –output =accumulation

The mass flow rate

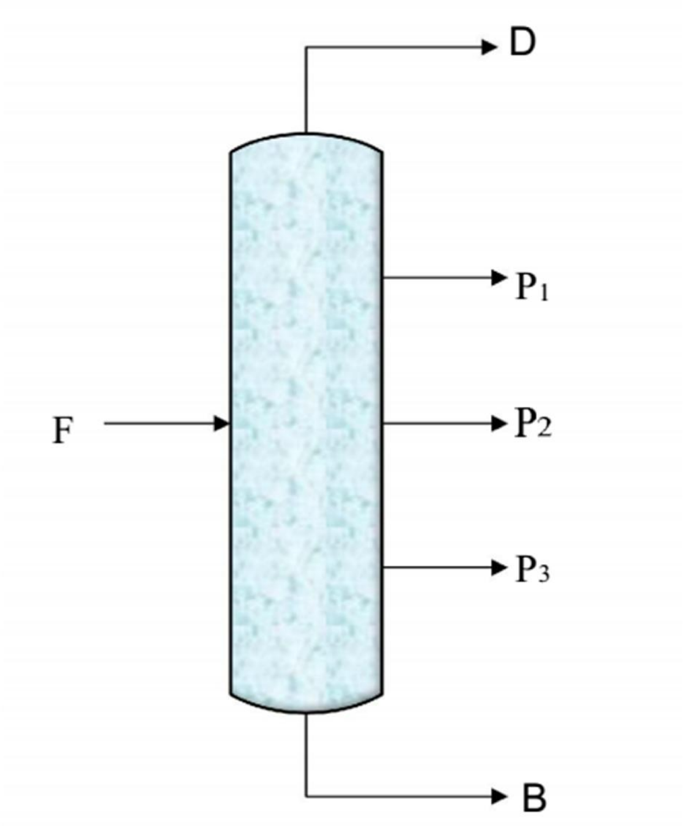


Figure 4.2 distillation tower

<u>At steady state</u>:-

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

Then:-

Input = output

For overall material balance:

$F = D + P_1 + P_2 + P_3 + B$

F: feed (crude oil)
D: gas + naphtha
P_1: kerosene
P_2: light diesel
P_3: heavy diesel
B: residue

Table 4.1 fractions and specific gravity of the products

Component	Concentration in feed	S.G
D	7.945	0.7313
P_1	3.873	0.8004
P_2	14.2480	0.8438
P_3	7.2480	0.83561
B	66.356	0.9220
Total	100%	

The mass flow rate of
D=258*0.07945=20.5 ton/hr. =4418334*0.07945=4225.02393 bbl/day

The mass flow rate of
P_1=258*0.03875=10 ton/hr =1882.o58 bbl/day

The mass flow rate of
P_2=258*0.14573=37.6 ton/hr =6716.1317 bbl/day

The mass flow rate of
P_3=258*0.07248=18.7 ton/hr =3372.939 bbl/day

The mass flow rate of
B =258*0.66356= 171.2 ton/hr =27986.186 bbl/day

Crude = (gas + naphtha) + kerosene + light diesel + heavy diesel + residue

Table 4.2: mass balance around distillation tower

Products	Cut mass %	Mass flow t/h	Mass flow Kg/h	Volume flow M3/h	Volume flow bbl/day	Density
Gas + naphtha	7.945	20.5	20500	28.03227	4225.02393	731.3
Kerosene	3.875	10	10000	12.493753	1882.0584	800.4
Light diesel	14.577	37.6	37600	44.56032	6716.1317	843.8
Heavy diesel	7.248	18.7	18700	22.3788	3372.939	835.61
Residue	66.356	171.2	171200	185.6832	27986.186	922
Crude (total)	100	258	258000	293.148	44183.34	880.1

4.2. Energy analysis

4.2.1. Streams heat flow calculation

The first law of thermodynamics can be written as fallow

Input of Energy - output Energy = accumulation of energy in process

We can assume

Steady state: accumulation = zero

rate of energy input = rate of energy output

Rate in = Σfi,in.(hi,in+ pei,in+ kei,in) + Q + W ……………………(4.1)

Rate out = Σfi,out.(hi,out + pei,out + kei,out) ……………………..(4.2)

fi = flow rate of component i,

hi = specific enthalpy,

pei = specific potential energy,

kei = specific kinetic energy

W = mechanical work done on system

Q = heat input to system

Changes in ke&pe smaller than enthalpy

$$\Sigma fi,out\ .hi,out - \Sigma fi,in\ .hi,in = Q + W \ldots\ldots\ldots\ldots..(4.3)$$

fi.hi = product of flow rate & specific enthalpy = total rate of energy transport with component i

Tabulated values of enthalpy are available only for the more common materials. In the absence of published data the following expressions can be used to estimate the specific enthalpy (enthalpy per unit mass). For pure materials, with no phase change.

$$H_T = \int_{T_d}^{T} C_P\ .dT \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ (4.4)$$

Where

H_T: specific enthalpy at temperature T,

Cp: specific heat capacity of the material, constant pressure,

T_d: the datum temperature.

Table (4.3): Streams heat flow

Stream	**Mw**	**mass flow (kg/h)**	**molar enthalpy(kj/kmol)**	**mass enthalpy(kj/kg)**	**heat flow(kj/h)**
Nile blend	260.5	258000	460853.8936	1769.112835	456431111.5
flash vapor	77.72160666	724.6691059	129081.6217	1660.820295	1203545.158
flash liquid	262.2616023	257275.3309	464007.2614	1769.253514	455185283.3
main steam	18.01510048	2294	228681.3603	12693.87093	29119739.92
light diesel steam	18.01510048	1294	231487.4195	12849.63244	16627424.38
heavy diesel steam	18.01510048	1294	231487.4195	12849.63244	16627424.38
off gas	61.32283296	511.0508808	140285.039	2287.647719	1169104.382
Naphtha	92.80291344	20499.93671	211710.8839	2281.295663	46766416.7
waste water	18.01510048	4841.954022	285534.7254	15849.7437	76743730.27
Kerosene	144.7112501	9529.534585	272746.3189	1884.762372	17960908.21
light diesel	196.7815403	37600.2539	365177.022	1855.748366	69776609.74
heavy diesel	252.2980153	18700.25749	430216.4389	1705.191531	31887520.7
Residue	396.0290597	171199.6545	584849.9363	1476.785408	252825151.6

4.2.2 Units operation energy analysis:

rate energy input = rate energy output

$Q_{in} = Q_{out}$

$Q = mCp\Delta T$ …………………………(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 = temperture change in C°

1. **Preheater**

258000 kg/h of Nile blend to be heated from 45 C° to 198.3 C° by exchanging with hot product’s stream.

We assume steady state: accumulation = zero

rate energy input = rate energy output

$Q_{in} = Q_{out}$

Heat load on preheater, Q = 2.255 * 258000*(198.3 – 45) = $8.91*10^{7}$kj/hr.

2. **Furnace duty**

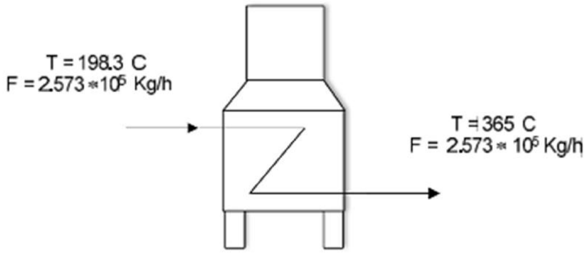


Figure 4.3: furnace input and output streams

We assume steady state: accumulation = zero

rate energy input = rate energy output

$Q_{in} = Q_{out}$

From HYSYS case
Cp =3.15 KJ/Kg.C□

duty $=2.573\text{x}10^{5}$ x 3.15 x (365-198.3) = $1.35\text{x}10^{8}$ kj/hr

3. **Pumps around**

A. **Heat removed by pump around_1**

Table 4.4: Pump around energy balance

Unit operation	Inlet temperature(C□)	Outlet temperature(C□)	Mass flow rate (Kg/h)
Pump around_ 1	116.4	44.1	9x104
Pump around_2	149.7	98	2.742x105
Pump around_3	248.6	100.6	10^{5} x 2.5

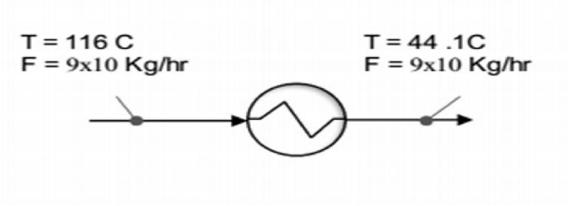


Figure 4.4: Pump around 1 input and output streams

$Q = mCp\Delta T$
$Cp = 2.1760$ KJ/Kg.C□
Q $=9\text{x}10^{4}$ x2.1760 x (116.4 - 44.1) $=1.415\text{x}10^{7}$Kj/hr

B. heat removed by pump around_2

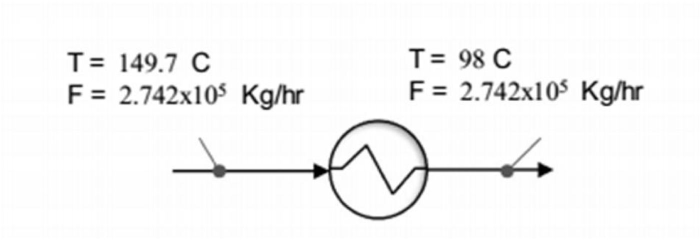


Figure 4.5: Pump around 2 input and output streams

$Q = mCp\Delta T$

$Cp = 2.334$ KJ/Kg.C□

Q $=2.742\text{x}10^{5}$ x2.334 x (149.7 -98) $=3.3\text{x}10^{7}$Kj/hr

C. heat removed by pump around_3

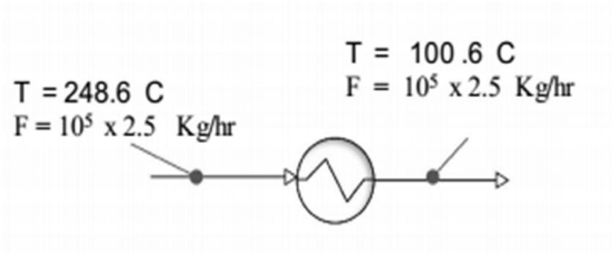


Figure 4.6: Pump around 3 input and output streams

$Q = mCp\Delta T$

$Cp = 2.5$ KJ/Kg.C□

Q = $10^{5\ \text{x}}$2.5x(248.6-100.6) = $3.7\text{x}10^{7}$Kj/hr

4. Condenser duty

Steady state   accumulation =zero

Energy In = energy out

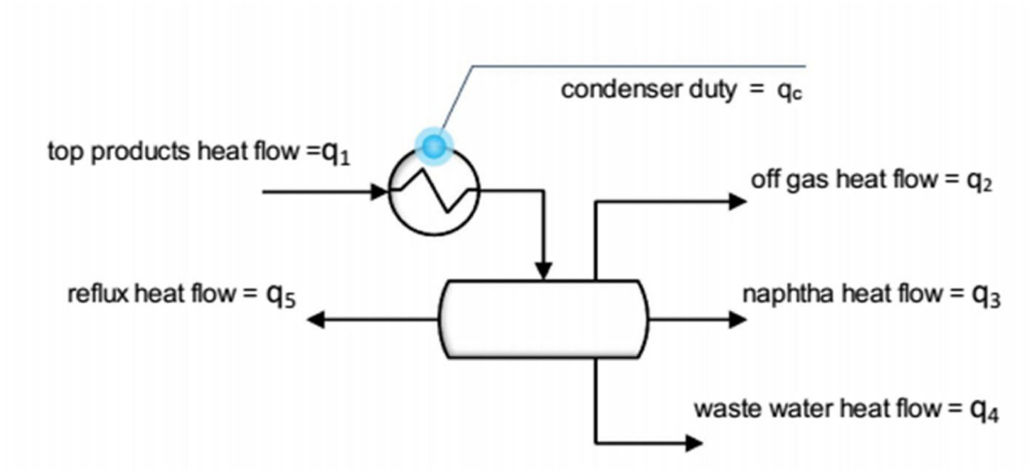


Figure 4.7: Condenser input and output streams

$$q_1 + q_c = q_2 + q_3 + q_4 + q_5$$

Table 4.5: Condenser streams heat flow

Stream	Heat flow (kj/hr)
q_1	$1.38 * 10^8$
q_2	$1.169 * 10^6$
q_3	$4.676 * 10^7$
q_4	$7.682 * 10^7$
q_5	$4.464 * 10^7$

Condenser duty = $3.139 * 10^7$ kj/hr

Table 4.6: Unit operation duty

Unit operation	Duty (Kj/hr)
Preheater	8.91×10^{7}
Furnace	1.35×10^{8}
Pump around 1	1.415×10^{7}
Pump around 2	3.3×10^{7}
Pump around 3	3.7×10^{7}

Summary of energy & material balances

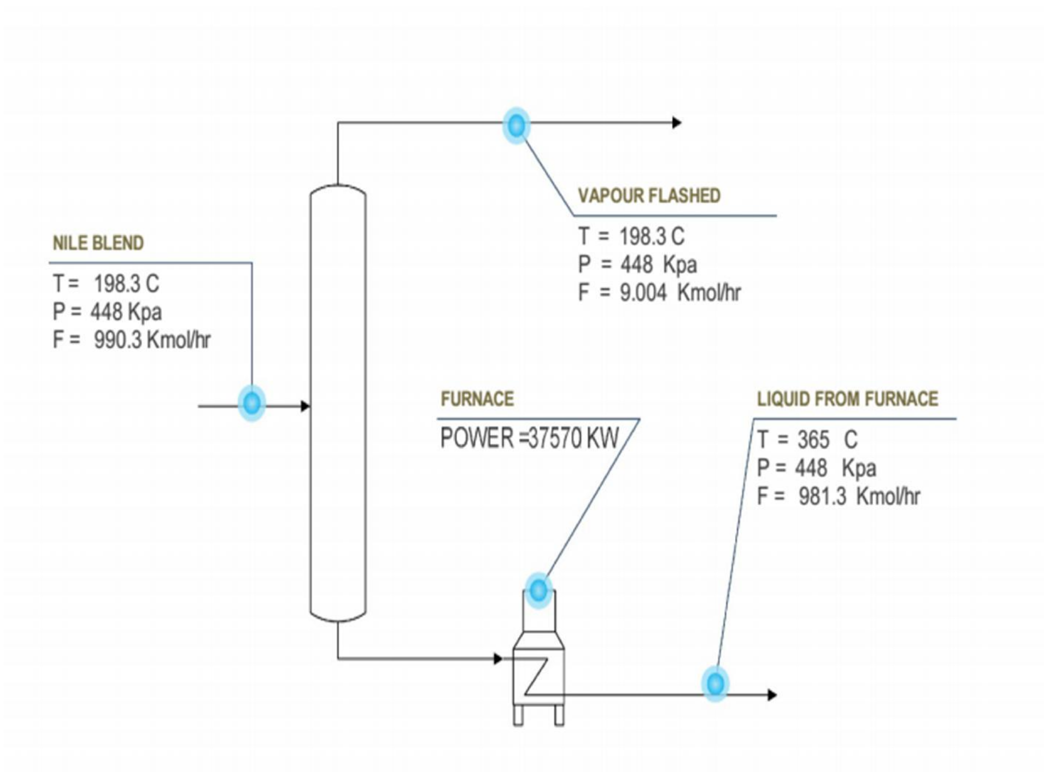


Figure 4.8: The Pre-Fractionation train

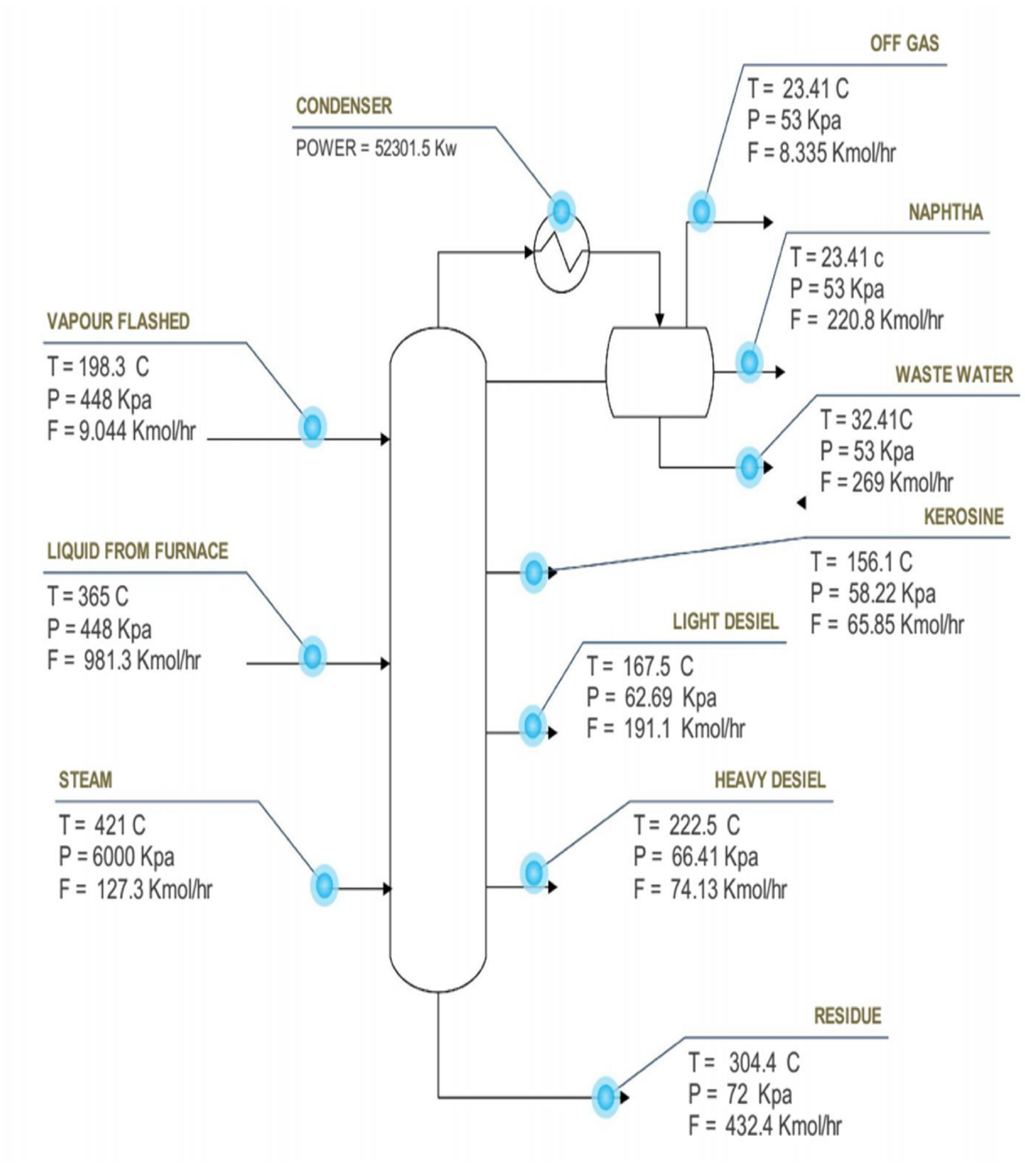


Figure 4 .9: Atmospheric distillation column

Chapter 5

Design of crude distillation tower

Designing steps of distillation column:

1. Calculation of Minimum number of stages N_{min}
2. Calculation of Minimum Reflux Ratio R_m.
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

5.1. Number of stages and total reflux

5.1.1. Minimum reflux and minimum number of stages

1 - Naphtha and kerosene

	Component	x_d	x_b
Light key	NBP[0]109*	0.000249255975689725	0.00186672956341614
Heavy key	NBP[0]643*	9.76381790508883E-09	0.0354673082887123

Fenske Equation: Minimum Number of theoretical trays

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

$$.\mathrm{Nmin} = \frac{\log(\frac{\mathrm{xD,l}}{\mathrm{xB,H}} \times \frac{\mathrm{xB,H}}{\mathrm{xB,l}})}{\log(\alpha \mathrm{l,H})} \ldots\ldots\ldots\ldots\ldots\ldots\ldots (5.1)$$

$\alpha l, H$is dependent on composition. For small variations in where$\alpha \mathrm{l, H} = \frac{\alpha\mathrm{,lk}}{\alpha\mathrm{,Hk}}$ is the relative volatility of the overhead vapor and $\alpha w$is the relative volatility of the Bottoms liquid

$$\alpha \mathrm{L, H} = \frac{\alpha\mathrm{,lk}}{\alpha\mathrm{,Hk}} = \frac{32.4983}{1} = 32.4983$$

$$\mathrm{Nmin} = \frac{\log\left(\frac{0.000249255975689725}{9.76381790508883E-09} \times \frac{0.0354673082887123}{0.00186672956341614}\right)}{\log(32.4983)}$$

$= \underline{\underline{3.76076}}$

Minimum Reflux ratio

Minimum reflux ratio (R_{min}) occurs when the intersecting point between the enriching line and q-line falls on VLE curve. Such a separation will require infinite steps/trays for distillation of the feed.

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q \qquad \text{\_..............(5.2)}$$

ϴ =1.04316

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_m + 1 \qquad \text{..............(5.3)}$$

Rmin+1=1.24687

Rmin = 0.24687

2- Kerosene and light diesel

	Component	x_d	x_b
Light key	NBP[0]473*	0.0136496841409066	0.0186368135581828
Heavy key	NBP[0]667*	2.63857531637522E-06	0.0332127395651005

Minimum number of stages

$$\alpha L, H = \frac{5.14}{1} = 5.14$$

$$\text{Nmin} = \frac{\log\left(\frac{0.0136496841409066}{2.63857531637522E-06} \times \frac{0.0332127395651005}{0.0186368135581828}\right)}{\log(5.14)}$$

Nmin = 5.576

Minimum reflux ratio

From equation 5.2

Θ =$\underline{\underline{0.7386}}$

From equation 5.3

Rmin+1=1.28916

Rmin = 0.28916

4- light diesel & heavy diesel :

	Component	Xd	Xb
Light key	NBP[0]618*	0.201104775824973	0.0393932327994476
Heavy key	NBP[0]1024*	4.15608477078005E-10	0.0161005925372012

Minimum number of stages

$$\alpha L, H = \frac{105}{1} = 105$$

$$Nmin = \frac{\log\left[\frac{0.201104775824973}{4.15608477078005E-10} * \frac{0.0161005925372012}{0.0393932327994476}\right]}{\log\,[105]}$$

$Nmin$ =$\underline{\underline{4.105}}$

Minimum reflux ratio

From equation 5.2

Θ =$\underline{\underline{1.22769}}$

From equation 5.3

Rmin +1 =1.00749

Rmin =0.007

5- Heavy diesel and residue

	Component	x_d	x_b
Light key	**NBP[0]618***	0.068052209077393	0.0393932327994476
Heavy key	NBP[0]1024*	3.34465628137494E-10	0.0161005925372012

Minimum number of stages

$$\alpha L, H = \frac{105}{1} = 105$$

$$\text{Nmin} = \frac{\log\left(\frac{xD,l}{xB,H} \times \frac{xB,H}{xB,l}\right)}{\log(\alpha l,H)}.$$

$$\text{Nmin} = \frac{log\left(\frac{0.068052209077393}{3.34465628137494E-10} \times \frac{0.0161005925372012}{0.0393932327994476}\right)}{\log(105)} = 3.919$$

Minimum reflux ratio

From equation 5.2

Ɵ =<u>3.3723</u>

From equation 5.3

Rmin+1=1.05656

Rmin = 0.05656

<u>Then :-</u>

Rmin = $\sum Rmin$=(0.24687)+(0.28916)+(0.00749)+(0.05656) =<u>0.60007</u>

$\text{Nmin} = \sum Nmin = 3.7607+5.57678+4.10522+3.91904= \underline{\underline{17.3618}}$

5.1.2. Actual Reflux Ratio

The rule of thumb is:

$R = (1.2 \text{ ------- } 1.5)\ R_{min}$

$R = 1.5\ R_{min}$

$R = 0.91211$

$$X = \frac{R-Rmin}{R+1} = \frac{0.91211-0.60007}{0.91211+1} = 0.16319$$

$$Y = (1 - X^{1/3}) = (1 - 0.16319^{1/3}) = 0.45353$$

5.1.3. Theoretical number of Plate

The Gilliland Correlation (1940) Used to calculate the number of stages, given reflux ratio, Rmin and Nmin

$$N = \frac{Nmin+Y}{1-Y} = \frac{17.3618+0.45353}{1-0.45353} = 32.6$$

5.1.4. Actual number of stage

$$\text{Actual number of stages=Na} = \frac{\text{Nth}}{\text{E}}$$

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

$$\textbf{Efficiency} = \mathbf{0.17-0.616\ log_{10}\,(\mu)}$$

Feed viscosity at average temperature $\mu f = 0.18$

$E = 0.17 - 0.616 * (\log 0.18) = 0.62$

$$N = \frac{32.6}{0.62} = 51.8\text{Tray}$$

5.2. Determination of the Column Diameter

5.2.1 Flow Parameter

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

V_n = mass vapor flow rate = 6.604e+004 kg/h

L_n= mass liquid flow rate = 1.912e+005 kg/h

Vapour density = 14.45 kg/m3

Liquid density = 642.6 kg/m3

$$F_{LV}=\left(\frac{1.912e+005}{6.604e+004}\right)\times\frac{14.45}{642.6}^{0.5}$$

F_{LV} = Liquid Vapor Factor = 0.434155

5.2.2Capacity Parameter

Assumed tray spacing = 18 inch (0.5 m)

From Fig (15-5) Plant Design and Economics for Chemical Engineering, sieve tray flooding capacity,

C_{sb} = 0.26 m/Sec

Surface tension of Mixture = σ = 8.151 dynes/Cm

Vapour density = 14.45 kg/m3

Liquid density = 642.6 kg/m3

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

V_{nf}=0.440231 m/sec

Assume 90% of flooding then

$V_n=0.9V_{nf}$

So, actual vapor velocity,

V_n=0.36208 m/sec

5.2.3. Daiameter of the tower

Net column area used in separation is

$$A_n = \frac{Qv}{Vn}$$

Volumetric flow rate of vapors = Q_v

$$Qv = \frac{\text{mass vapor flow rate}}{3600 \times \text{vapor density}}$$

Mass vapor flow rate = 6.604e+004 kg/h

Vapor density= 14.45 kg/m3

$$Qv = \frac{6.604e+004}{3600 \times 14.45}$$

$$Qv = 1.2695 \text{ m}^3/\text{sec}$$

Now, net area

$$A_n = \frac{Qv}{Vn} = \frac{1.2695}{0.396208} = 3.204155 \text{ m2}$$

Assume that downcommer occupies 15% of cross sectional Area (A_c) of column thus:

$$A_c = A_n + A_d$$

Where,

A_d = downcommer area

$$A_c = A_n + 0.15(A_c)$$

$$A_c = A_n / 0.85$$

$$A_c = \frac{3.6959}{0.85}$$

$$A_c = 3.76959 \text{ m}^2$$

So Diameter of Column Is

$$A_c = (\pi/4)D^2$$

$$D^{\wedge}2 = (4A_c/\pi)$$

$$D = 2.191 \text{ meter}$$

5.3 Height of Distillation Column

Height of column

$$H_c = (N_{act} - 1)\ H_s + \Delta H + \text{plates thickness}$$

No. of plates = 51.8

Tray Spacing: Hs = 0.46 to 0.61 m (0.3 and 0.91 m are also used)

$H_s = 0.50$ m

$\Delta H = 0.5$ meter each for liquid hold up and vapor disengagement

$$\Delta H = 1 \text{ m}$$

Total thickness of trays = 0.005*51.8= 0.259 m

So,

Height of column = (51.8 - 1)*0.50+ 1+0.259

= <u>26.6</u> meters

5.4. Provisional Plate Design:

Column Diameter D_c= 2.191 m

Column Cross-sectional Area(A_c)= 3.679 m^2

Down comer area Ad= $0.15A_c$ = 0.565 m^2

Net Area (A_n) = $A_c - A_d$ = 3.204 m^2

Active area $A_a = A_c - 2A_d$ = 2.549 m^2

Hole area A_h take 10% A_a = 0.1 × 2.549 = 0.2549 m^2

Weir length

Ad / Ac = 0.565/ 3.679 = 0.153

From Figure 5.1

L_w / dc = 0.82

L_w = 2.191*0.82

= 1.796 m

Weir length should be 60 to 85% of column diameter which is satisfactory

Take weir height, h_w= 50 mm

Hole diameter, d_h = 5 mm

Plate thickness = 5 mm

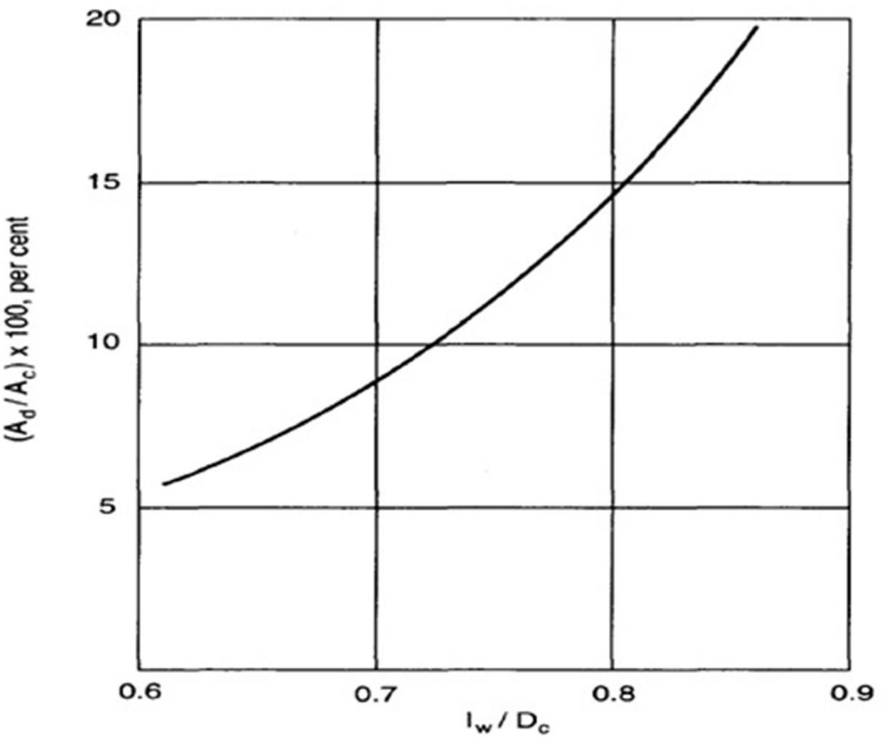


Figure 5.1: Relation between downcomer area and weir length

Coulson & Richardson 6th volume 3rd edition

Number of hole

Area of 1 Hole = $(\pi/4)\ D_{hole}^2$

D_{hole} = 5 mm

Area of 1 Hole = $(\pi/4)\ (0.005)^2 = 0.0000196\ m^2$

Area of N Holes = $0.1158\ m^2$

So,

Number OF Holes = $\frac{\text{Area of N Holes}}{\text{Area of 1 Hole}}$ = 5900

5.5. Liquid flow arrangement

In order to find liquid flow arrangement first find maximum liquid volumetric flow rate

So liquid flow rate = (Liquid mass rate)/ (3600) (Liquid density)

Lm = 1.912e+004/ (3600) (642.6)

Lm =0.0082

Max Liquid Rate Is At the bottom of column so using "L_m" values

So Maximum liquid flow rate = $0.0082\ m^3/sec$

Cross flow single pass plate is selected

From figure 5.2

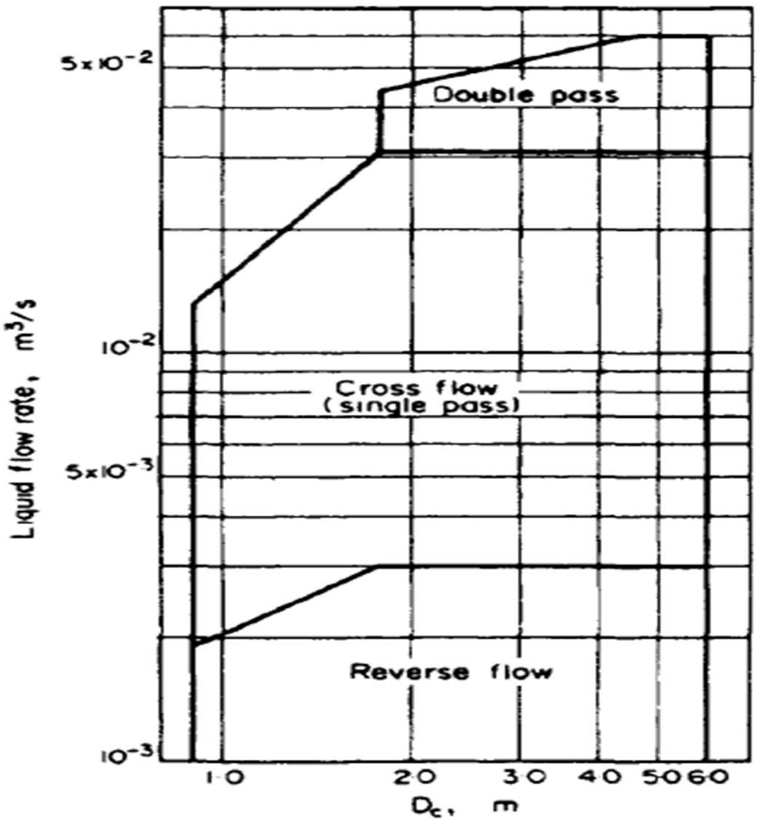


Figure 5.2: Flow passes Coulson & Richardson 6th volume 3rd edition

Chapter 6
Control system

6.1. Control of flash tower
Level control – LIC 101

The objective of this loop controller is to control the liquid present level in the vessel using direct action in the operation so the process variable PV is the vessel or the tower and the output target OP is the outlet liquid stream from flash tower , the set point SP here is set to be 50% of the vessel high so less than this present the the indicter will send a signal to the valve to control the level Inside the vessel as shown in figure (6.1) .

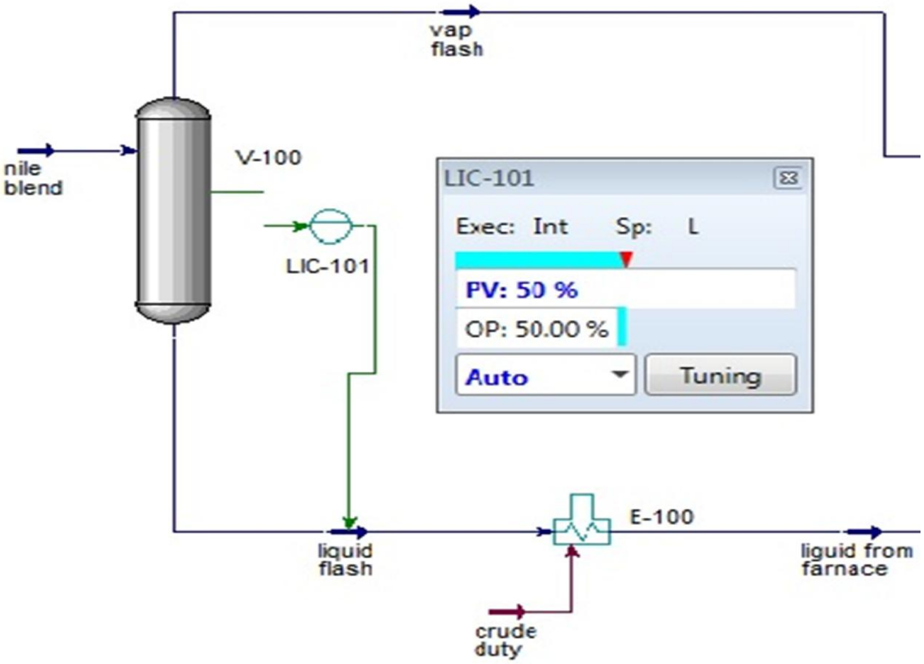


Figure 6.1: Level controls – LIC 101

6.2. Control of light diesel flow
Flow control – FIC 101

In this controller the objective is to control the light diesel flow that outlet from side stripper and the action of the control is direct, the variable here is the liquid volume flow and the process variable here PV is the light diesel stream and the outlet target OP is the light diesel draw from the tower and the set point is SP is set to maintain the flow to be 45.32 m3 h with a maximum PV 100 m3 h as shown in figure (6.2).

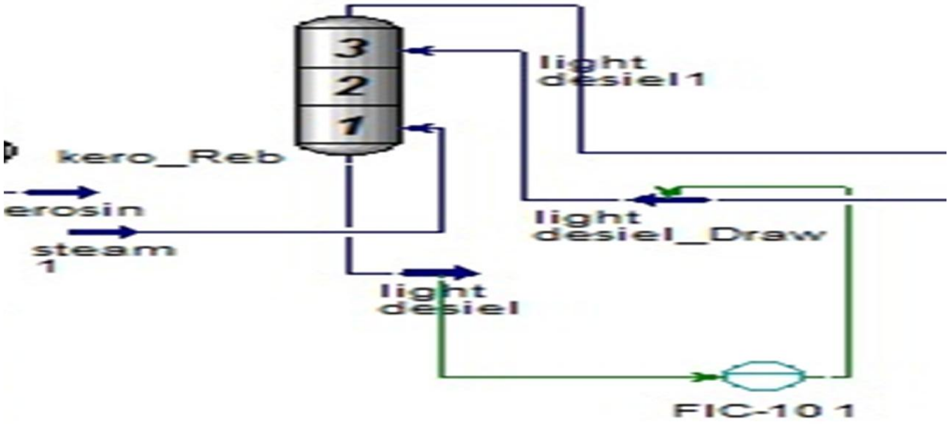


Figure 6.2: Flow control – FIC 101

6.3. Control of heavy diesel flow

Flow control – FIC 100

In this controller the objective is to control the heavy diesel flow that outlet from side stripper and the action of the control is direct, the variable here is the liquid volume flow and the process variable here PV is the heavy diesel stream and the outlet target OP is the heavy diesel draw from the tower and the set point is SP is set to maintain the flow to be 21.76 m3 h with a maximum PV 60 m3 h

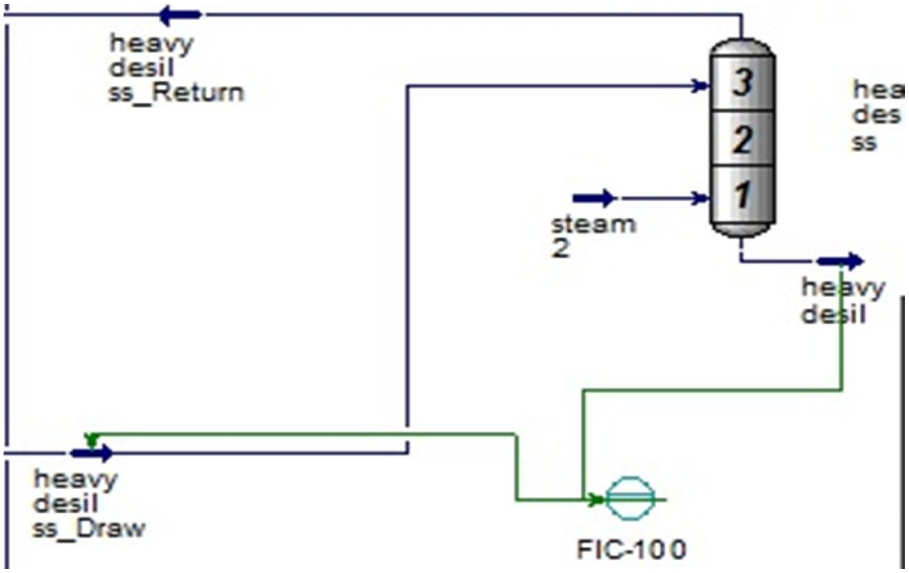


Figure 6.3: Flow control – FIC 100

6.4. Control kerosene reboiler

Level control – LIC 101

The objective of this loop controller is to control the liquid present level in the vessel using direct action in the operation so the process variable PV is the kerosene reboiler and the output target OP is the kerosene draw from the tower , the set point SP here is set to be 50% of the vessel high so less than this present the indicter will send a signal to the valve to control the level Inside the vessel as shown in figure (6.4) .

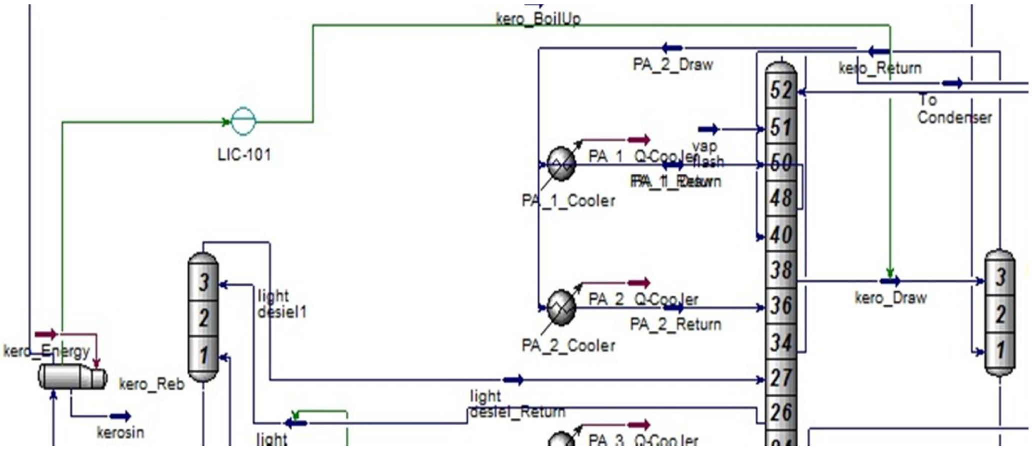


Figure 6.4: Level control – LIC 101

6.5. Control of off gas

Flow control – FIC 102

In this controller the objective is to control the off gas flow that outlet from condenser and the action of the control is reverse, the variable here is the liquid molar flow and the process variable here PV is the off gas stream and the outlet target OP is the duty of the condenser and the set point is SP is set to maintain the flow to be 8.334 kgmole/h with a maximum PV 511 kgmole/h as shown in figure (6.5)

6.6 Control condenser

Level control – LIC 100

The objective of this loop controller is to control the liquid present level in the condenser using direct action in the operation so the process variable PV is the condenser and the output target OP is the reflux from the condenser, the set point SP here is set to be 50% of the condenser high so less than this Present the indicter will send a signal to the valve to control the level inside the condenser as shown in figure (6.5).

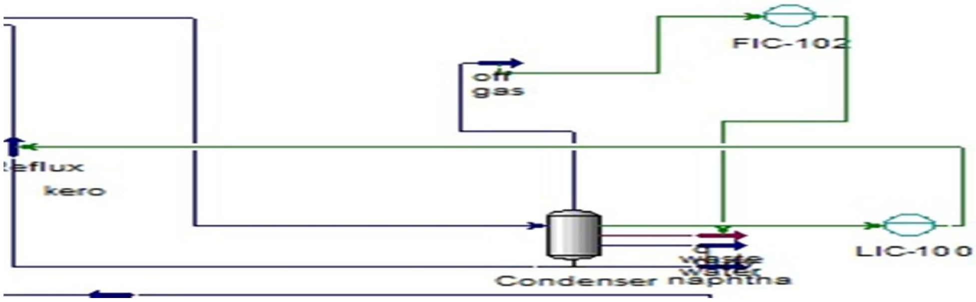
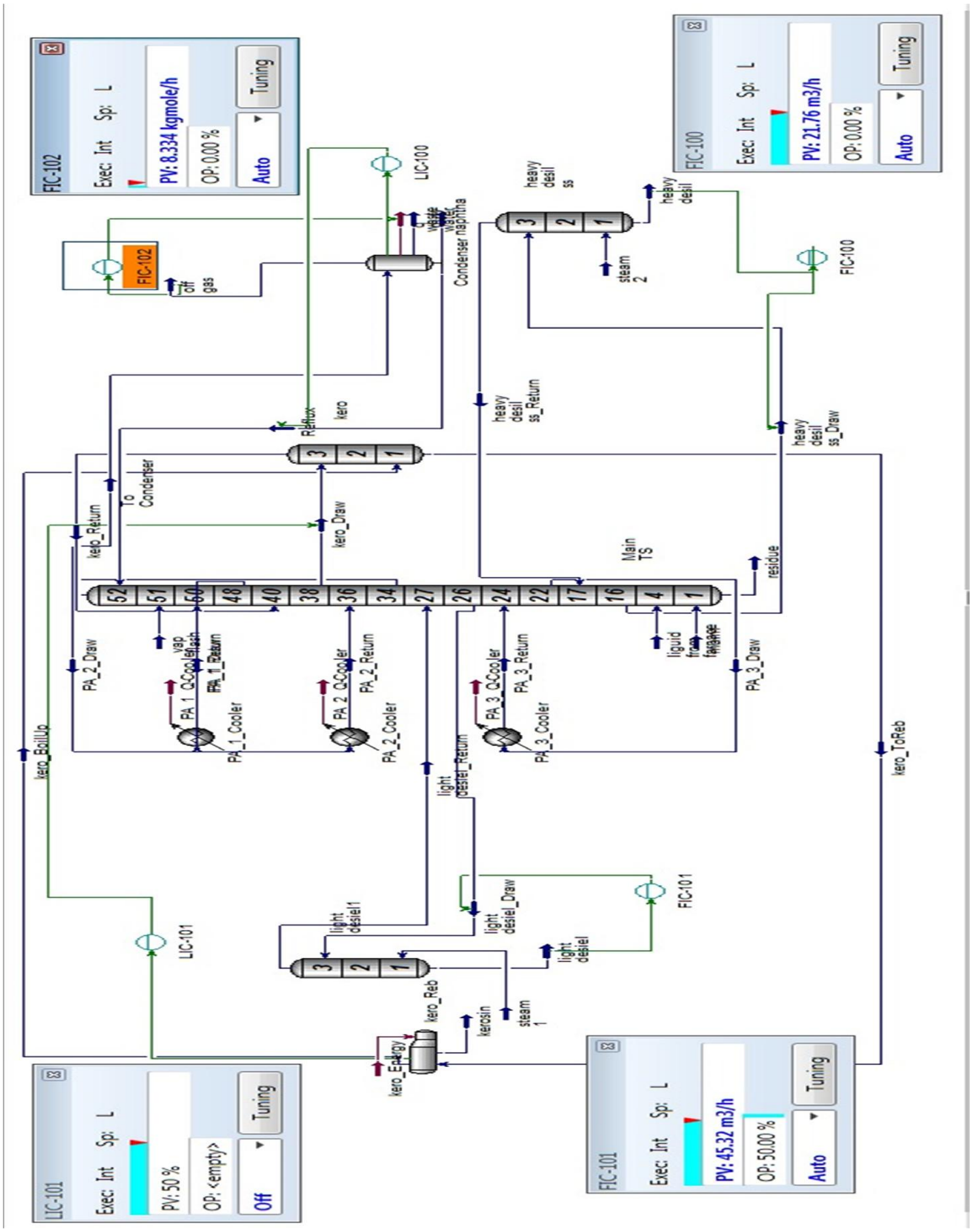


Figure 6.5: Level control – LIC 100 and Flow control – FIC 102



FIC-102
Exec: Int Sp: L
PV: 8.334 kgmole/h
OP: 0.00 %
Auto
Tuning

FIC-100
Exec: Int Sp: L
PV: 21.76 m3/h
OP: 0.00 %
Auto
Tuning

LIC-101
Exec: Int Sp: L
PV: 50 %
OP: <empty>
Off
Tuning

FIC-101
Exec: Int Sp: L
PV: 45.32 m3/h
OP: 50.00 %
Auto
Tuning

Figure 6.6: Control loops of the distillation process

Chapter 7
Cost estimation

7.1. Equipment cost

1. **Distillation tower**

$C = F_1.C_b+N.F_2.F_3.F_4.C_t+C_p$..(7.1)

$C_b= \exp[7.123 + 0.11478(\ln W) + 0.02488(\ln W)^2 + 0.0158\left(\frac{L}{D}\right)\ln(\left(\frac{Tb}{Tp}\right)$…(7.2)

D=7.19 ft
L = 87.27ft
N = 52 sieve trays
Wall thickness = 0.50 in. for pressure,
Tb = 0.75 in. at the bottom,
Flanged and dished heads weigh 325 Ib each

$$W=\frac{3.14}{4}\times\left(\frac{16\times 87.27\times 0.5\times 501}{12}\right)+2\times 325 = 23531.321\text{lb}$$

$C_b= \exp[7.123 + 0.11478(\ln 32.12) + 0.02488(\ln 32129)^2 + 0.0158\left(\frac{120}{4}\right)\ln\left(\frac{0.75}{0.5}\right)$
C_b=54524.54

$F_1 = 1.7$
$F_2= 1.189 + 0.0577(7.19) = 1.604$
$F_3 = 0.85$
$F_4= 1$
$C_t= 375.8 \exp[0.1739(D)]=375.8 \exp[0.1739(7.19)] = 1312.11$

$Cp = 2049(D^{0.6332}\times L^{0.8}) = 2049(7.19^{0.6332}\times 87.27^{0.8}) = 255102.5$

Purchase price

C = 1.7(54524.54) + 52(1.604)(0.85)(1312.11)+ 255102.5 =$440818.6

From Table C.1 , the installation factor is 2.1 so that the installed
Price is

C installed= 2.1(440818.6)= $881637.2

2. **Heat exchangers**

Shell-and-tube :

$C = F_d.\ F_m.\ F_p.\ C_b$...(7.3)
price in $

$C_{b} = \exp[8.821 - 0.30863(\ln A) + 0.0681(\ln A^{2})]$………(7.4)

Area = 900 ft^2

F_d= exp [-1.1156+0.0906(lnA)] =exp [-1.1156+0.0906(ln 150)] = 0.516006

F_b = 0.7771 + 0.04981(1n A) =0.7771 + 0.04981(1n 150) = 1.02668

$F_m = g_1 + g_2$ (In A)………………………………..(7.5)

g_1= 0.8603 g_2 = 0.23296

F_m= 0.8603 + 0.23296 (In 150) = 2.02757

C_b =exp [8.821 -0.30863(ln 150) +0.0681(ln 150^2)] = 7977.002

Purchase price:

$C = F_d. F_m. F_p. C_b$ = 8568.546 $

Installed Cost = (purchase price)*multiplier = 8568.546 * 2.2 = 18850.8 $

For 16 heat exchanger = 16 *18850.8 = 301612.8 $

3. Pumps

Centrifugal pump prices

$C = F_m. F_p. C_b$

C_b $= 1.55 \exp[8.833 - 0.6019(\ln Q\sqrt{H}) + 0.0519 \left(\ln Q\sqrt{H}^{2}\right)$ ………………….(7.6)

Material - Stainless steel, 304

F_m =9.7

ft=exp $\left[b1 + b2(\ln Q\sqrt{H}) + b3 \left(\ln Q\sqrt{H}^{2}\right)\right]$ ………………………………….(7.7)

b1 =5.1029
b2 = -1.2217
b3 = 0.0771
Q =2100 gpm H =150 ft. HP = 75

Cb =32.52445
F_b =1.9105

Purchase price

C= 602.65 $

Installed Cost = (purchase price)*multiplier = 602.65 * 2 = 1205.31 \$
For 15 pumps = 15 * 1205.31 =18079.67 \$

4. Fired heater (furnace):

$$C= (1+F_{D+}F_p)Q^{0.88} \text{ ………………………..(7.8)}$$

Table 7.1: Furnace cost required

Tube material	K
CrMo steel	33.8
Design type	Fd
Process heater	0.1
Design pressure psi	Fp
3000	0.6
Duty	40 million btu/hr

A box type fired heater for crude oil at1500 psig with a duty of 40 million Btu/hr. the installed price is:

$C_{insulated}$= 3 3.8 (1.0 + 0.10 + 0.15) $(40)^{0.85}$ = \$1,008,320.

5. Stripper

$$C = F_1.C_b+N.F_2.F_3.F_4.C_t+C_p \text{……………………..(7.9)}$$

C_b= $\exp[7.123 + 0.11478(\ln W) + 0.02488(\ln W)^2 + 0.0158\left(\frac{L}{D}\right)\ln(\left(\frac{Tb}{Tp}\right). (7.10)$
D= 4.92 ft,
L = 70ft,
N = 9 sieve trays,
Wall thickness 6 = 0.50 in. for pressure,
Tb = 0.75 in. at the bottom,
W=32129

C_b= $\exp[7.123 + 0.11478(\ln 32{,}12) + 0.02488(\ln 32129)^2 + 0.0158\left(\frac{120}{4}\right)\ln\left(\frac{0.75}{0.5}\right)$
C_b=17659.19
F_1= 1.7,
F_2= 1.189 + 0.0577(4.93) = 1.47
F_3 = 0.85
F_4= 1
C_t = 375.8 exp [0.1739(D)] = 375.8 exp [0.1739(4.92)] = 884.16

C_b =$2049(D^{0.6332} \times L^{0.8})$ =$2049(4.92^{0.6332} \times 70^{0.8})$= 168179.35

Purchase price C = 1.7(17659.17) + 9(1.47) (0.85) (884.16) +168179.35 =\$208142.8

From Table C.1 , the installation factor is 2.1 so that the installed
Price is

C installed= 2.1(208142.8) = $416285.6

Equipments cost = cost of tower + cost of heat exchanger + cost of pumps + cost of furnace + cost of stripper

Equipments cost = 881637.2+301612.8 +18079.67 +1008320 +416285=$ 2625934.6

7.2. Operating labor and maintenance costs:-

1. **Maintenance cost**

Maintenance		
Cost/8000 Hours		379000$
Total Maintenance Cost	Cost/period	415005$

2. **Supervision**

Supervisors per Shift		1
Unit Cost	Cost/Supervisor/H	35$
Total Supervision Cost	Cost/period	306600$

3. **Steam & electricity**

Description	Rate	Units	Rate Units	Cost per Hour	Cost /period
Electricity	67.24	KW	KW	5.2111	45649.236$
Steam	2.270538	KLB	KLB/H	18.482179	161903.888$

Chapter 8

Results and discussion

8.1. Results from design:-

Specification Sheet of Distillation Column:

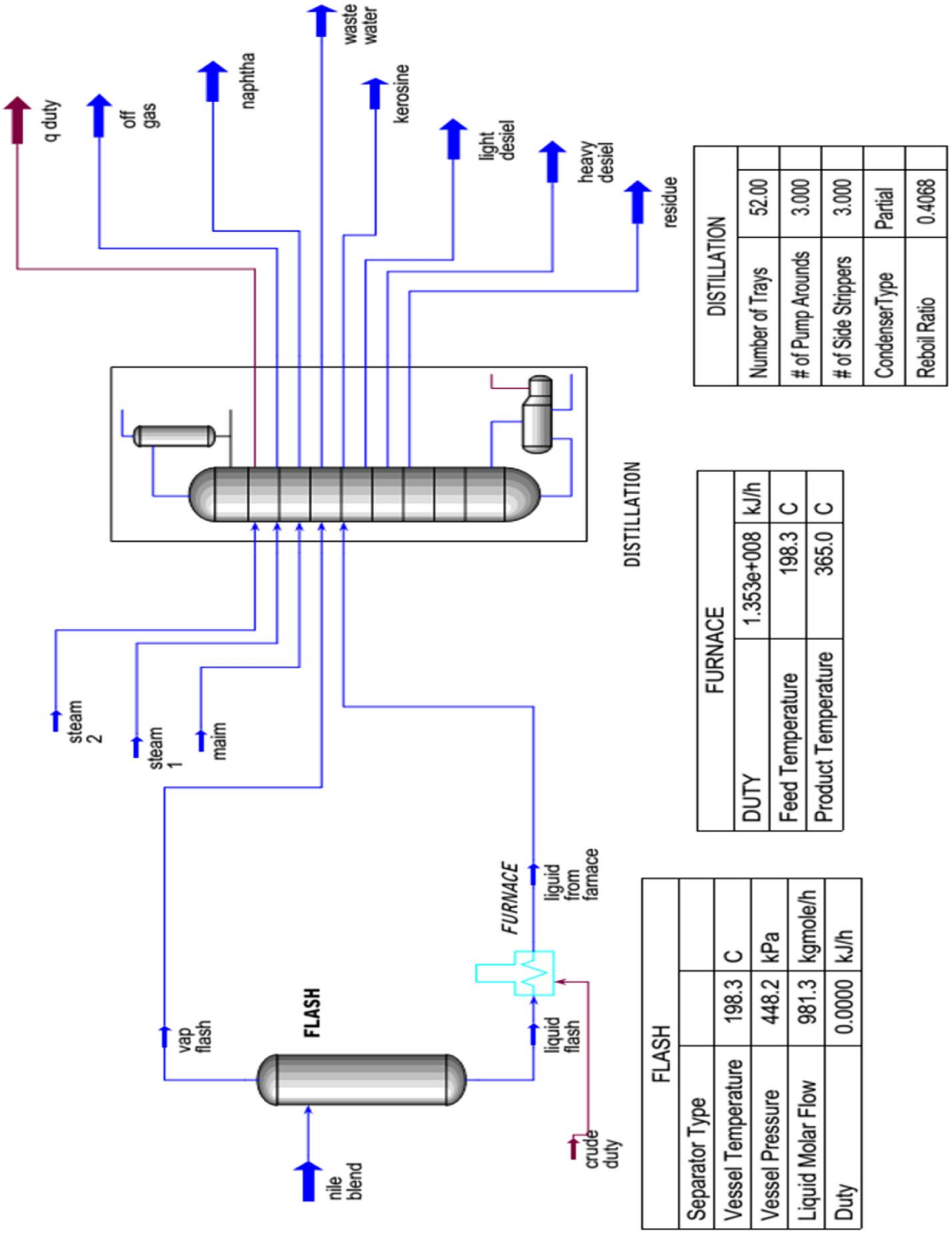
Identification:

Item :	Distillation column
No. required:	1
Tray type:	Sieve tray
Function:	Separation of **Nile blend**
Operation:	Continuous

Design results:

Table 8.1: Design results

No. of tray	52	Active holes	5900
Pressure	0.053MPa (Top) 0.072MPa(bottom)	Tray thickness	0.5 m
Height of column	26.6 m	Weir height	50 mm
Diameter of column	2.191 m	Reflux ratio	0.91211
Hole size	5 mm	Tray spacing	0.5 m
Weir length	1.796 m	Active area	2.549m^2

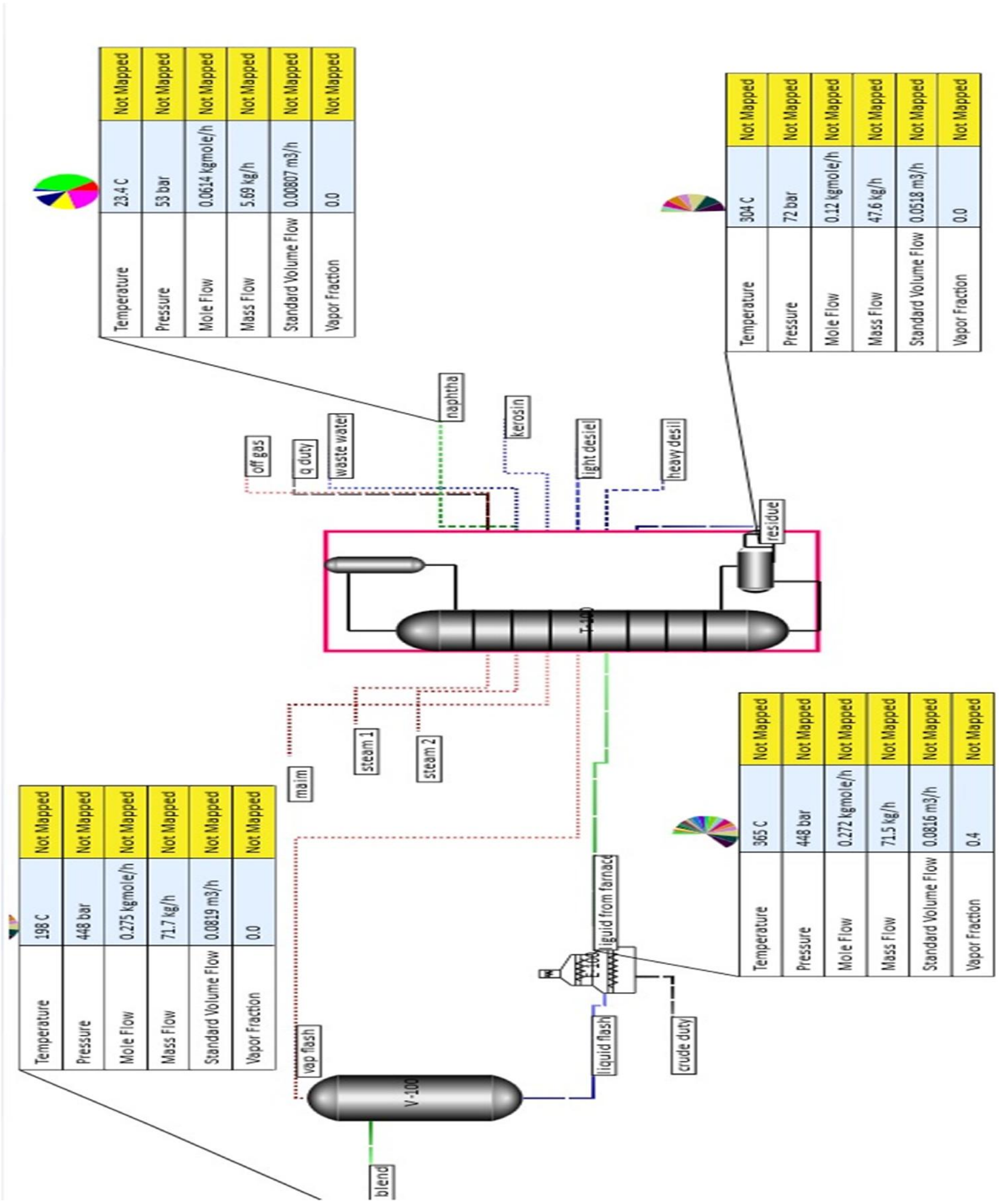


FLASH		
Separator Type		
Vessel Temperature	198.3	C
Vessel Pressure	448.2	kPa
Liquid Molar Flow	981.3	kgmole/h
Duty	0.0000	kJ/h

FURNACE		
DUTY	1.353e+008	kJ/h
Feed Temperature	198.3	C
Product Temperature	365.0	C

DISTILLATION	
Number of Trays	52.00
# of Pump Arounds	3.000
# of Side Strippers	3.000
CondenserType	Partial
Reboil Ratio	0.4068

Figure 8.1 : Simulation flow sheet



Temperature	23.4 C	Not Mapped
Pressure	53 bar	Not Mapped
Mole Flow	0.0614 kgmole/h	Not Mapped
Mass Flow	5.69 kg/h	Not Mapped
Standard Volume Flow	0.00807 m3/h	Not Mapped
Vapor Fraction	0.0	Not Mapped

Temperature	304 C	Not Mapped
Pressure	72 bar	Not Mapped
Mole Flow	0.12 kgmole/h	Not Mapped
Mass Flow	47.6 kg/h	Not Mapped
Standard Volume Flow	0.0518 m3/h	Not Mapped
Vapor Fraction	0.0	Not Mapped

Temperature	198 C	Not Mapped
Pressure	448 bar	Not Mapped
Mole Flow	0.275 kgmole/h	Not Mapped
Mass Flow	71.7 kg/h	Not Mapped
Standard Volume Flow	0.0819 m3/h	Not Mapped
Vapor Fraction	0.0	Not Mapped

Temperature	365 C	Not Mapped
Pressure	448 bar	Not Mapped
Mole Flow	0.272 kgmole/h	Not Mapped
Mass Flow	71.5 kg/h	Not Mapped
Standard Volume Flow	0.0816 m3/h	Not Mapped
Vapor Fraction	0.4	Not Mapped

Figure 8.2: HYSYS plant view of crude distillation unit

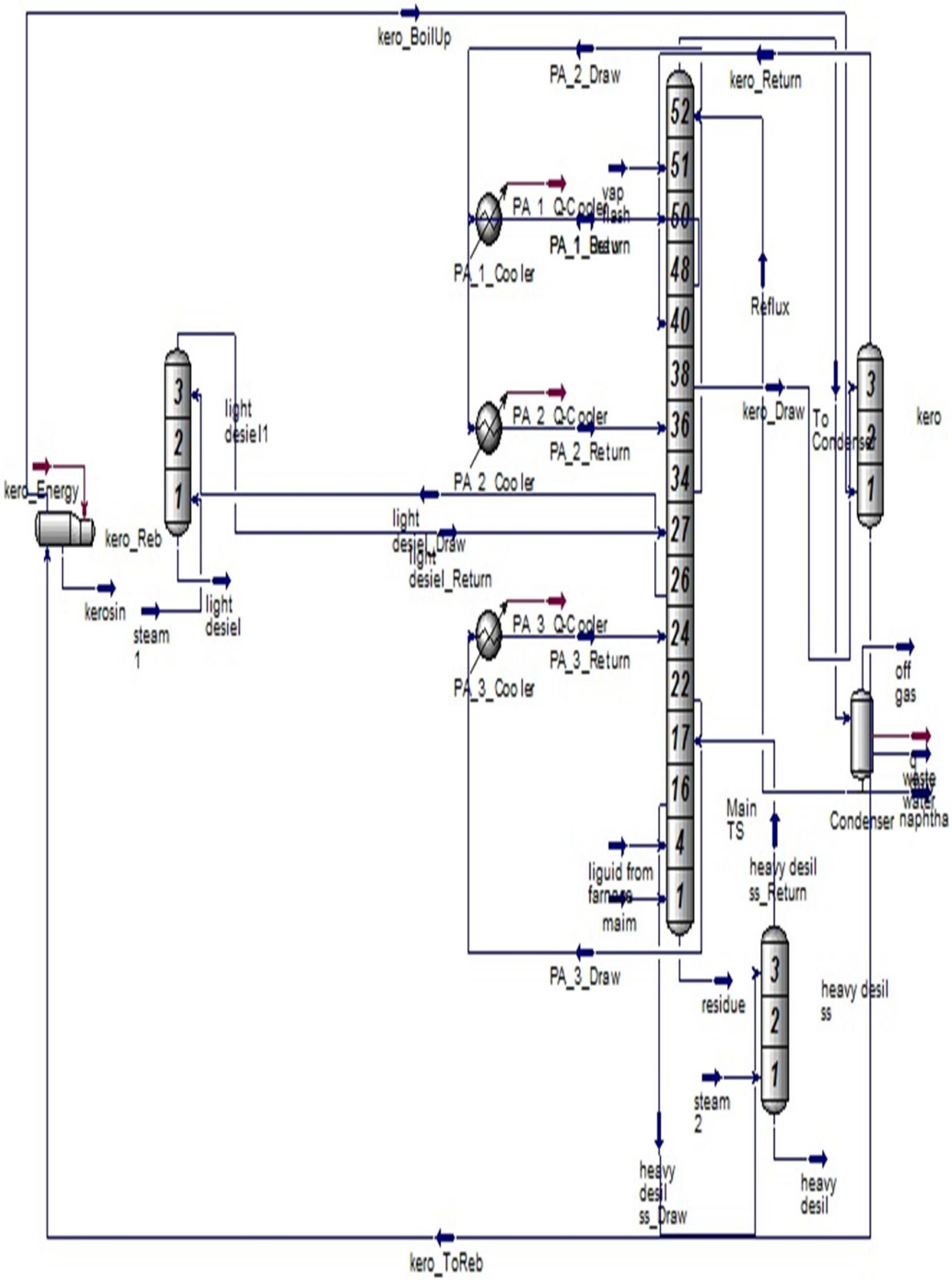
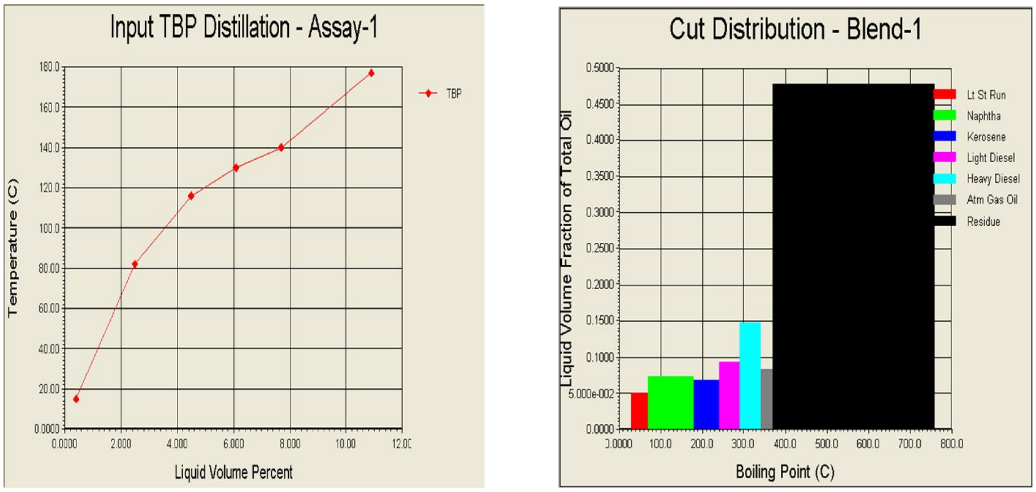


Figure 8.3: Column environment

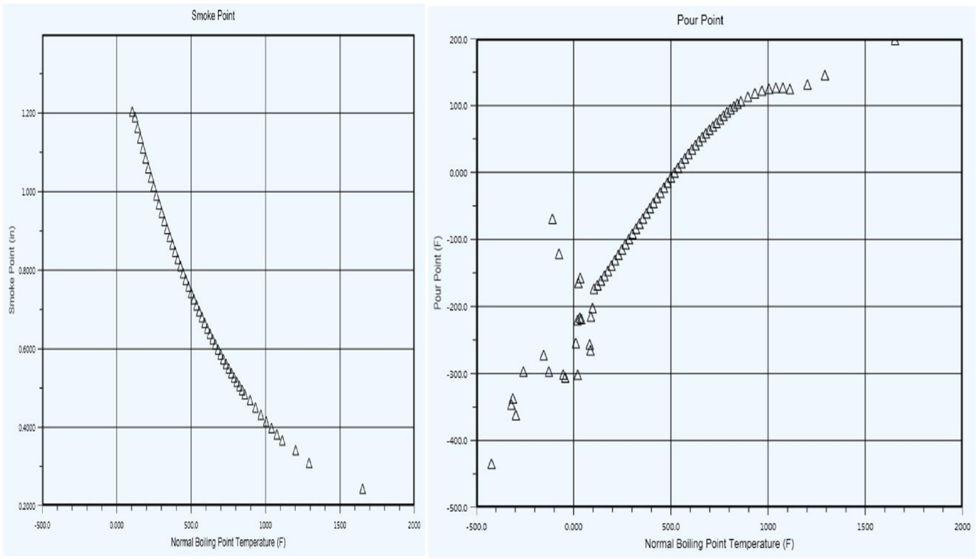
8.2 Curves results

1- Nile blend properties


(a)

(b)

Figure 8.4: (a) TBP Distillation (b) Cut distribution


(c)

(d)

Figure 8.5 : (c) Smoke point (d) Pour point

2- Colum profiles

A. Pressure profile

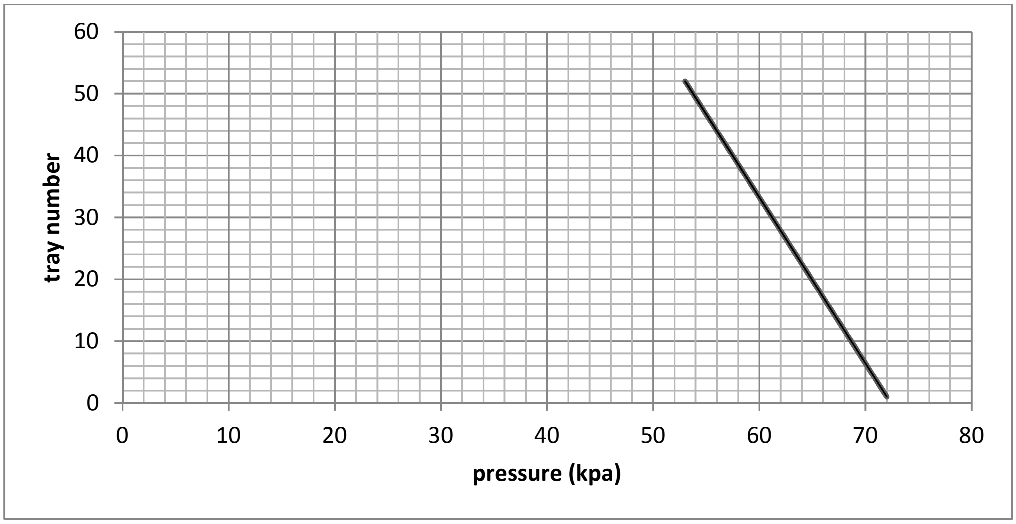


Figure 8.6: Pressure profile

B. Temperature profile

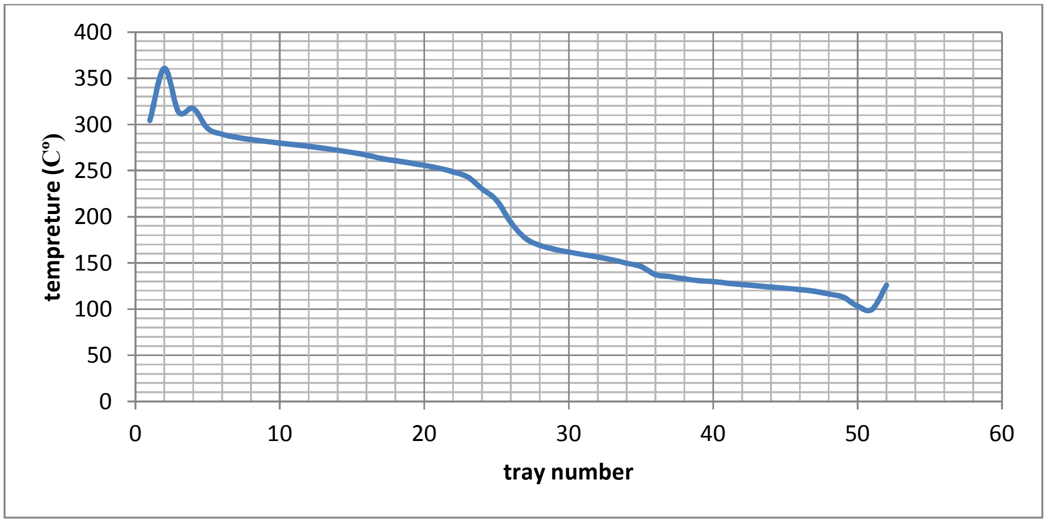


Figure 8.7 : Temperature profile

C. Net Flow profile

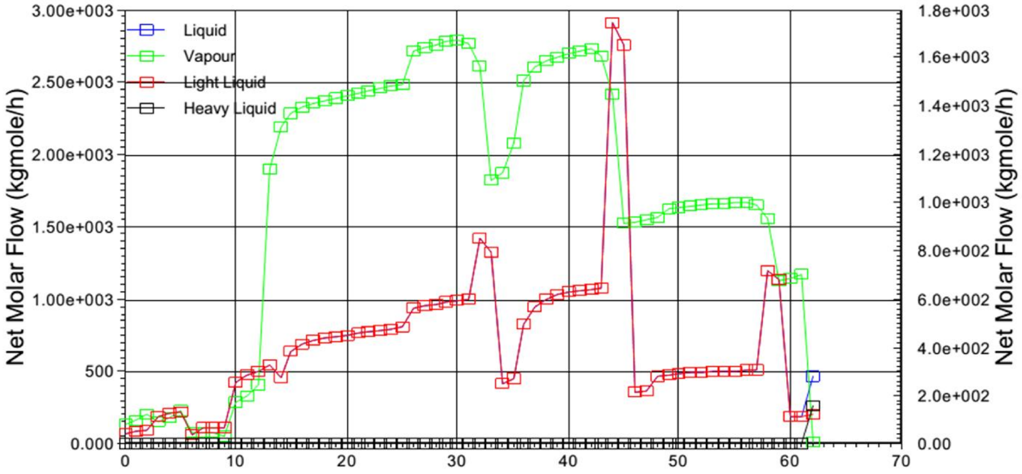


Figure 8.8: Net flow profiles

D. Properties profile

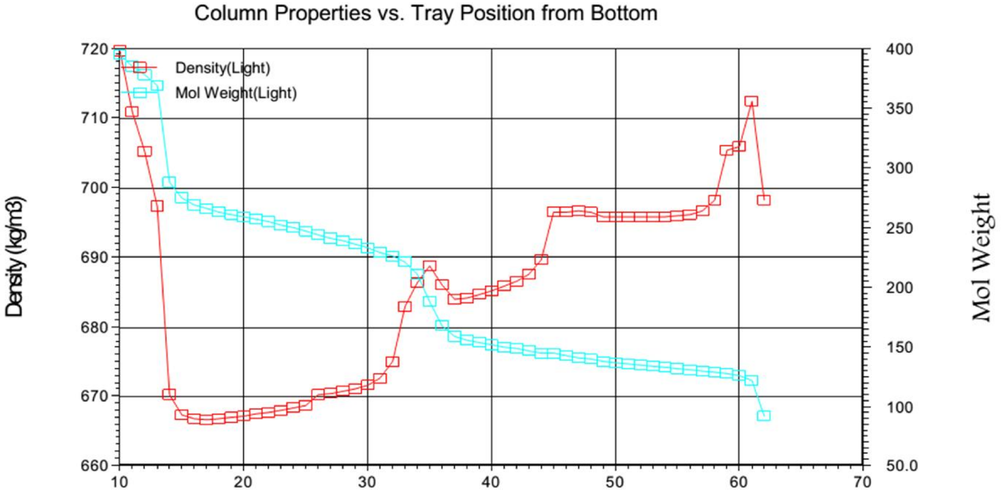


Figure 8.6: column properties profiles

8.3. Trays dimension and efficiency

Table 8.2: Trays dimensions and efficiency

Main fractionation column tray dimensions						
tray number	internal type	tray spacing	Diameter	weir height	weir length	Efficiency
52__Main TS	Sieve	0.5	1.5	50	1.2	0.62
51__Main TS	Sieve	0.5	1.5	50	1.2	0.62
50__Main TS	Sieve	0.5	1.5	50	1.2	0.62
49__Main TS	Sieve	0.5	1.5	50	1.2	0.62
48__Main TS	Sieve	0.5	1.5	50	1.2	0.62
47__Main TS	Sieve	0.5	1.5	50	1.2	0.62
46__Main TS	Sieve	0.5	1.5	50	1.2	0.62
45__Main TS	Sieve	0.5	1.5	50	1.2	0.62
44__Main TS	Sieve	0.5	1.5	50	1.2	0.62
43__Main TS	Sieve	0.5	1.5	50	1.2	0.62
42__Main TS	Sieve	0.5	1.5	50	1.2	0.62
41__Main TS	Sieve	0.5	1.5	50	1.2	0.62
40__Main TS	Sieve	0.5	1.5	50	1.2	0.62
39__Main TS	Sieve	0.5	1.5	50	1.2	0.62
38__Main TS	Sieve	0.5	1.5	50	1.2	0.62
37__Main TS	Sieve	0.5	1.5	50	1.2	0.62
36__Main TS	Sieve	0.5	1.5	50	1.2	0.62
35__Main TS	Sieve	0.5	1.5	50	1.2	0.62
34__Main TS	Sieve	0.5	1.5	50	1.2	0.62
33__Main TS	Sieve	0.5	1.5	50	1.2	0.62
32__Main TS	Sieve	0.5	1.5	50	1.2	0.62
31__Main TS	Sieve	0.5	1.5	50	1.2	0.62
30__Main TS	Sieve	0.5	1.5	50	1.2	0.62
29__Main TS	Sieve	0.5	1.5	50	1.2	0.62
28__Main TS	Sieve	0.5	1.5	50	1.2	0.62
27__Main TS	Sieve	0.5	1.5	50	1.2	0.62
26__Main TS	Sieve	0.5	1.5	50	1.2	0.62
25__Main TS	Sieve	0.5	1.5	50	1.2	0.62
24__Main TS	Sieve	0.5	1.5	50	1.2	0.62
23__Main TS	Sieve	0.5	1.5	50	1.2	0.62
22__Main TS	Sieve	0.5	1.5	50	1.2	0.62
21__Main TS	Sieve	0.5	1.5	50	1.2	0.62
20__Main TS	Sieve	0.5	1.5	50	1.2	0.62
19__Main TS	Sieve	0.5	1.5	50	1.2	0.62
18__Main TS	Sieve	0.5	1.5	50	1.2	0.62
17__Main TS	Sieve	0.5	1.5	50	1.2	0.62
16__Main TS	Sieve	0.5	1.5	50	1.2	0.62
15__Main TS	Sieve	0.5	1.5	50	1.2	0.62
14__Main TS	Sieve	0.5	1.5	50	1.2	0.62
13__Main TS	Sieve	0.5	1.5	50	1.2	0.62

12__Main TS	Sieve	0.5	1.5	50	1.2	0.62
11__Main TS	Sieve	0.5	1.5	50	1.2	0.62
10__Main TS	Sieve	0.5	1.5	50	1.2	0.62
9__Main TS	Sieve	0.5	1.5	50	1.2	0.62
8__Main TS	Sieve	0.5	1.5	50	1.2	0.62
7__Main TS	Sieve	0.5	1.5	50	1.2	0.62
6__Main TS	Sieve	0.5	1.5	50	1.2	0.62
5__Main TS	Sieve	0.5	1.5	50	1.2	0.62
4__Main TS	Sieve	0.5	1.5	50	1.2	0.62
3__Main TS	Sieve	0.5	1.5	50	1.2	0.62
2__Main TS	Sieve	0.5	1.5	50	1.2	0.62
1__Main TS	Sieve	0.5	1.5	50	1.2	0.62

8.4. Comparison between HYSYS results and experiment results:-

Experiments been in the central petroleum laboratories (CPL) type of gas chromatography (GC) was performed on a sample of petroleum products produced from crude distillation unit(CDU) to be utilized in the process of simulation and also compare the results of laboratory experiments with the results provided by the program HYSYS .

The comparison between the results from laboratory experiments and the results of HYSYS in the following form:-

1. **For naphtha:-**

Table 8.3: Naphtha comparison

property name	Results from experimental data	Results from HYSYS analysis
Sp gr	0.715	0.7142
Mw	102.738	93
Riedvapor p (psi)	6.84	10.65
Riedvapor p (kpa)	47.18	82.2
Temperature	66.5	66.5
Composition		
methane	1.00E-03	1.00E-04
Ethane	0.00232	0.0011
propane	0.01238	0.0019
Butane	0.0299	0.0426
pentane	0.04377	0.4909

2. **For kerosene:**

Table 8.4: kerosene comparison

property name	from experimental data	from HYSYS analysis
Sp gr	0.774	0.775
Mw	154.681	144
Riedvapor p (psi)	0.1	0.065
Riedvapor p (kpa)	0.66	0.44
Temperature	120.38	120
Composition		
C10	0.0003	0.0001
C11	0.0058	0.002
C12	0.02779	0.0099
C13	0.32	0.0242
C4	0.25	0.1805

8.5 Comparison between costs:

As a result of the outcome of the program calculates the total cost of the detail alone depending on world prices and some constants and so we calculate the total cost of computational methods alone and in detail to compare it with the result that provided by the program are as follows :

Table 8.5: Cost comparison

Equipment name	Equipment type	Cost calculated	Cost from HYSYS
Distillation	Tower	881637.2	3567200
Heat exchanger	H E X	56552.4	46100
Pumps	Pump	1205.31	1216.67
Furnace	Heater	1008320	63400
Stripping	Stripping tower	416285	160300

8.6 Control of the naphtha and kerosene production

1. Naphtha

Table 8.6: properties of Naphtha before increasing

Naphtha before increasing	
Flow rate	20000 kg/hr
Mass percentage	7.752%
Density @15 C°	705.9 kg/m^3
Molecular weight	93
Kinematic viscosity @23.41 C°	0.5505 (cSt)

Table 8.7: properties of Naphtha after increasing

Naphtha after increasing	
Flow rate	30000 kg/hr
Mass percentage	11.628%
Density @15 C°	734.8 kg/m^3
Molecular weight	104.6
Kinematic viscosity @23.41 C°	0.6395 (cSt)

2. Kerosene

Table 8.8: properties of kerosene before increasing

kerosene before increasing	
Flow rate	10000 kg/hr
Mass percentage	3.876%
Density @15 C°	790.2 kg/m^3
Molecular weight	144.7
Kinematic viscosity @165.1C°	0.3838 (cSt)

Table 8.9: properties of kerosene after increasing

kerosene a after increasing	
Flow rate	15000 kg/hr
Mass percentage	5.814%
Density @15 C°	794.3 kg/m^3
Molecular weight	149.6
Kinematic viscosity @165.1C°	0.3830 (cSt)

Conclusion and recommendation

Conclusion

The simulation have been performed using specifications such as products flow rate and duties of condenser ,pump arounds and reboiler .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 Khartoum Refinery Company KRC to reach 52 trays and diameter 2m and 26.6m, and then control the unit using a different type of controller and performed a method to increasing the production of naphtha and kerosene in term of change on the monitor of the HYSYS software , where the results of increasing of naphtha from 7.75% to 11.62% , and for kerosene from 3.87%to 5.8% . The research used the program to estimate the cost of equipment and the operation cost of the unit and also used hand calculation to determine the cost of the unit.

Recommendation

We recommend to use HYSYS program in every project in the future and used the new version of the program in the laboratory because it’s very useful software and to teaches advances course of the program beside the basic course.

APPENDICES

Appendix A

Simulation data from Khartoum Refinery Company

Table A.1 Flash feed operating conditions

Temperature	198.3 c
Flow rate	258 t/h

Table A.2 Stripper specification

Stripper	type	Draw stage	Return stage
Kerosene	Reboiler	38	40
Light diesel	Steam	26	27
Heavy diesel	Steam	16	17

Table A.3 Distillation column operating conditions

	DRAW	RETURNE	DRAW	$T_{1\,(C)}$	$T_{2\,(C)}$

TOP STAGE	[illegible]	[illegible]
BOTTOM STAGE	360.7ºC	0.072 Mpa

	STAGE	STAGE	RATE(t/h)		
PA-1	48	50	20	150.4	78.16
PA-2	34	36	75	192.5	140.8
PA-3	22	24	40	301.5	153.5

Table A.3 Pumps abound operating conditions

Table A.4 production data

Naphtha	20.5 t/h
Kerosene	10 t/h
Light diesel	37.6 t/h
Heavy diesel	18.7 t/h
Residue	192 t/h

Appendix B

HYSYS reports

Column Sub-Flowsheet: T-100 @Main

CONDITIONS

Name		maim @Main	from farnace @Main	vap flash @Main	steam 1 @Main	steam 2 @Main
Vapour		1.0000	0.4286	1.0000	1.0000	1.0000
Temperature	(C)	421.0000 *	365.0000 *	198.3000	300.0000 *	300.0000 *
Pressure	(kPa)	6000.0000 *	448.1593	448.1593	100.0000 *	99.9740 *
Molar Flow	(kgmole/h)	127.3376	980.9874	9.3239	71.8286	71.6083
Mass Flow	(kg/h)	2294.0000 *	257275.3309	724.6691	1294.0000 *	1290.0299 *
Std Ideal Liq Vol Flow	(m3/h)	2.2986	293.6251	1.1145	1.2966	1.2926
Molar Enthalpy	(kJ/kgmole)	-2.287e+005	-3.261e+005	-1.291e+005	-2.315e+005	-2.315e+005
Molar Entropy	(kJ/kgmole-C)	167.7	868.9	234.1	196.4	196.4
Heat Flow	(kJ/h)	-2.9120e+07	-3.1994e+08	-1.2035e+06	-1.6627e+07	-1.6576e+07
Name		**residue @Main**	**off gas @Main**	**naphtha @Main**	**waste water @Main**	**kerosin @Main**
Vapour		0.0000	1.0000	0.0000	0.0000	0.0000
Temperature	(C)	304.3490	23.3602	23.3602	23.3602	156.0850
Pressure	(kPa)	72.0000	53.0000	53.0000	53.0000	58.2157
Molar Flow	(kgmole/h)	432.2918	8.3334	220.8982	268.5926	65.8527
Mass Flow	(kg/h)	171199.9632	511.0118	20500.0027	4838.7223	9529.6185
Std Ideal Liq Vol Flow	(m3/h)	185.4389	0.8492	29.3443	4.8485	12.0611
Molar Enthalpy	(kJ/kgmole)	-5.848e+005	-1.403e+005	-2.117e+005	-2.855e+005	-2.727e+005
Molar Entropy	(kJ/kgmole-C)	1195	175.3	82.01	53.28	219.3
Heat Flow	(kJ/h)	-2.5283e+08	-1.1691e+06	-4.6767e+07	-7.6693e+07	-1.7961e+07
Name		**light desiel @Main**	**heavy desil @Main**	**q duty @Main**		
Vapour		0.0000	0.0000	---		
Temperature	(C)	167.4867	222.5078	---		
Pressure	(kPa)	62.6863	66.4118	---		
Molar Flow	(kgmole/h)	191.0751	74.1189	---		
Mass Flow	(kg/h)	37600.0707	18700.0255	---		
Std Ideal Liq Vol Flow	(m3/h)	45.3243	21.7627	---		
Molar Enthalpy	(kJ/kgmole)	-3.652e+005	-4.302e+005	---		
Molar Entropy	(kJ/kgmole-C)	373.2	577.6	---		
Heat Flow	(kJ/h)	-6.9776e+07	-3.1887e+07	3.1272e+07		

PROPERTIES

Name		maim @Main	guid from farnace @Mai	vap flash @Main	steam 1 @Main	steam 2 @Main
Molecular Weight		18.02	262.3	77.72	18.02	18.02
Molar Density	(kgmole/m3)	1.135	0.2016	0.1214	2.104e-002	2.103e-002
Mass Density	(kg/m3)	20.45	52.86	9.438	0.3790	0.3789
Act. Volume Flow	(m3/h)	112.2	4867	76.78	3414	3405
Mass Enthalpy	(kJ/kg)	-1.269e+004	-1244	-1661	-1.285e+004	-1.285e+004
Mass Entropy	(kJ/kg-C)	9.307	3.313	3.012	10.90	10.90
Heat Capacity	(kJ/kgmole-C)	42.65	802.7	187.7	36.27	36.27
Mass Heat Capacity	(kJ/kg-C)	2.368	3.061	2.416	2.013	2.013
LHV Molar Basis (Std)	(kJ/kgmole)	0.0000	---	---	0.0000	0.0000
LHV Mass Basis (Std)	(kJ/kg)	---	---	---	---	---
Phase Fraction [Vol. Basis]		1.000	0.2823	1.000	1.000	1.000
Phase Fraction [Mass Basis]		1.000	0.2567	1.000	1.000	1.000
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)	112.2	4569	76.78	3414	3405
Avg. Liq. Density	(kgmole/m3)	55.40	3.341	8.366	55.40	55.40
Specific Heat	(kJ/kgmole-C)	42.65	802.7	187.7	36.27	36.27

Column Sub-Flowsheet: T-100 @Main (continued)

PROPERTIES

Name		maim @Main	guid from farnace @Mai	vap flash @Main	steam 1 @Main	steam 2 @Main
Act. Liq. Flow	(m3/s)	---	8.267e-002	---	---	---
Z Factor		0.9160	---	---	0.9975	0.9975
Watson K		---	11.65	12.79	---	---
User Property		---	---	---	---	---
Cp/(Cp - R)		1.242	1.010	1.046	1.297	1.297
Cp/Cv		1.414	1.007	1.060	1.301	1.301
Heat of Vap.	(kJ/kgmole)	2.876e+004	3.262e+005	5.463e+004	4.103e+004	4.103e+004
Kinematic Viscosity	(cSt)	1.268	---	1.150	53.43	53.44
Liq. Mass Density (Std. Cond)	(kg/m3)	1015	876.2	658.9	1015	1015
Liq. Vol. Flow (Std. Cond)	(m3/h)	2.261	293.8	1.101	1.276	1.272
Liquid Fraction		0.0000	0.5714	0.0000	0.0000	0.0000
Molar Volume	(m3/kgmole)	0.8811	4.961	8.235	47.53	47.55
Mass Heat of Vap.	(kJ/kg)	1596	1244	702.9	2277	2277
Phase Fraction [Molar Basis]		1.0000	0.4286	1.0000	1.0000	1.0000
Surface Tension	(dyne/cm)	---	8.151	---	---	---
Thermal Conductivity	(W/m-K)	6.379e-002	---	3.100e-002	4.335e-002	4.335e-002
Viscosity	(cP)	2.592e-002	---	1.086e-002	2.025e-002	2.025e-002
Cv (Semi-Ideal)	(kJ/kgmole-C)	34.34	794.3	179.4	27.96	27.96
Mass Cv (Semi-Ideal)	(kJ/kg-C)	1.906	3.029	2.309	1.552	1.552
Cv	(kJ/kgmole-C)	30.15	796.8	177.1	27.87	27.87
Mass Cv	(kJ/kg-C)	1.674	3.038	2.278	1.547	1.547
Cv (Ent. Method)	(kJ/kgmole-C)	---	---	---	---	---
Mass Cv (Ent. Method)	(kJ/kg-C)	---	---	---	---	---
Cp/Cv (Ent. Method)		---	---	---	---	---
Reid VP at 37.8 C	(kPa)	6.467	20.43	239.5	6.467	6.467
True VP at 37.8 C	(kPa)	6.467	29.31	528.2	6.467	6.467
Liq. Vol. Flow - Sum(Std. Cond)	(m3/h)	2.261	294.1	1.100	1.275	1.271
HHV Molar Basis (Std)	(kJ/kgmole)	4.101e+004	---	---	4.101e+004	4.101e+004
HHV Mass Basis (Std)	(kJ/kg)	2276	---	---	2276	2276
CO2 Loading		---	---	---	---	---
CO2 Apparent Mole Conc.	(kgmole/m3)	---	---	---	---	---
CO2 Apparent Wt. Conc.	(kgmol/kg)	---	---	---	---	---
Phase Fraction [Act. Vol. Basis]		1.000	0.9389	1.000	1.000	1.000
Mass Exergy	(kJ/kg)	1263	328.6	149.4	632.2	632.1
Partial Pressure of H2S	(kPa)	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity Index		6.327	---	-10.01	---	---

Name		residue @Main	off gas @Main	naphtha @Main	waste water @Main	kerosin @Main
Molecular Weight		396.0	61.32	92.80	18.02	144.7
Molar Density	(kgmole/m3)	1.814	2.186e-002	7.518	55.98	4.689
Mass Density	(kg/m3)	718.5	1.340	697.7	1009	678.6
Act. Volume Flow	(m3/h)	238.3	381.3	29.38	4.798	14.04
Mass Enthalpy	(kJ/kg)	-1477	-2288	-2281	-1.585e+004	-1885
Mass Entropy	(kJ/kg-C)	3.017	2.858	0.8837	2.957	1.515
Heat Capacity	(kJ/kgmole-C)	1143	103.4	191.0	77.71	363.7
Mass Heat Capacity	(kJ/kg-C)	2.885	1.687	2.058	4.314	2.513
LHV Molar Basis (Std)	(kJ/kgmole)	---	---	---	8.804e-005	---
LHV Mass Basis (Std)	(kJ/kg)	---	---	---	4.887e-006	---
Phase Fraction [Vol. Basis]		0.0000	1.000	0.0000	0.0000	0.0000
Phase Fraction [Mass Basis]		0.0000	1.000	0.0000	0.0000	0.0000
Partial Pressure of CO2	(kPa)	0.0000	0.0000	0.0000	0.0000	0.0000

Column Sub-Flowsheet: T-100 @Main (continued)

PROPERTIES

Name	residue @Main	off gas @Main	naphtha @Main	waste water @Main	kerosin @Main
Act. Gas Flow (ACT_m3/h)	---	381.3	---	---	---
Avg. Liq. Density (kgmole/m3)	2.331	9.813	7.528	55.40	5.460
Specific Heat (kJ/kgmole-C)	1143	103.4	191.0	77.71	363.7
Std. Gas Flow (STD_m3/h)	1.022e+004	197.0	5223	6351	1557
Std. Ideal Liq. Mass Density (kg/m3)	923.2	601.8	698.6	998.0	790.1
Act. Liq. Flow (m3/s)	6.618e-002	---	8.162e-003	1.333e-003	3.901e-003
Z Factor	8.265e-003	0.9836	2.860e-003	3.840e-004	3.479e-003
Watson K	11.75	13.38	12.33	19.50	11.79
User Property	---	---	---	---	---
Cp/(Cp - R)	1.007	1.087	1.046	1.120	1.023
Cp/Cv	1.132	1.093	1.267	1.146	1.116
Heat of Vap. (kJ/kgmole)	3.346e+005	4.244e+004	4.568e+004	4.181e+004	4.343e+004
Kinematic Viscosity (cSt)	0.3575	5.253	0.5502	0.9169	0.3838
Liq. Mass Density (Std. Cond) (kg/m3)	918.7	612.9	705.2	1015	788.9
Liq. Vol. Flow (Std. Cond) (m3/h)	186.3	0.8337	29.07	4.768	12.08
Liquid Fraction	1.000	0.0000	1.000	1.000	1.000
Molar Volume (m3/kgmole)	0.5512	45.75	0.1330	1.786e-002	0.2132
Mass Heat of Vap. (kJ/kg)	844.8	692.0	492.2	2321	300.1
Phase Fraction [Molar Basis]	0.0000	1.0000	0.0000	0.0000	0.0000
Surface Tension (dyne/cm)	13.61	---	19.23	72.38	14.23
Thermal Conductivity (W/m-K)	9.473e-002	1.513e-002	0.1087	0.6086	0.1003
Viscosity (cP)	0.2569	7.040e-003	0.3838	0.9247	0.2605
Cv (Semi-Ideal) (kJ/kgmole-C)	1134	95.12	182.7	69.40	355.4
Mass Cv (Semi-Ideal) (kJ/kg-C)	2.864	1.551	1.969	3.852	2.456
Cv (kJ/kgmole-C)	1009	94.65	150.7	67.81	326.0
Mass Cv (kJ/kg-C)	2.549	1.543	1.624	3.764	2.253
Cv (Ent. Method) (kJ/kgmole-C)	1041	---	157.2	64.21	---
Mass Cv (Ent. Method) (kJ/kg-C)	2.628	---	1.694	3.564	---
Cp/Cv (Ent. Method)	1.098	---	1.215	1.210	---
Reid VP at 37.8 C (kPa)	8.411e-005	562.4	73.43	6.468	0.4504
True VP at 37.8 C (kPa)	6.470	1297	82.28	6.468	0.4504
Liq. Vol. Flow - Sum(Std. Cond) (m3/h)	186.3	0.8337	29.07	4.768	12.08
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)	---	---	---	---	---
Phase Fraction [Act. Vol. Basis]	0.0000	1.000	0.0000	0.0000	0.0000
Mass Exergy (kJ/kg)	206.5	-24.23	-5.970e-002	-3.742e-002	51.38
Partial Pressure of H2S (kPa)	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity Index	-16.97	-19.86	-6.514	2.032	-14.89
Name	**light desiel @Main**	**heavy desil @Main**			
Molecular Weight	196.8	252.3			
Molar Density (kgmole/m3)	3.638	2.806			
Mass Density (kg/m3)	715.8	708.0			
Act. Volume Flow (m3/h)	52.53	26.41			
Mass Enthalpy (kJ/kg)	-1856	-1705			
Mass Entropy (kJ/kg-C)	1.896	2.289			
Heat Capacity (kJ/kgmole-C)	491.8	673.7			
Mass Heat Capacity (kJ/kg-C)	2.499	2.670			

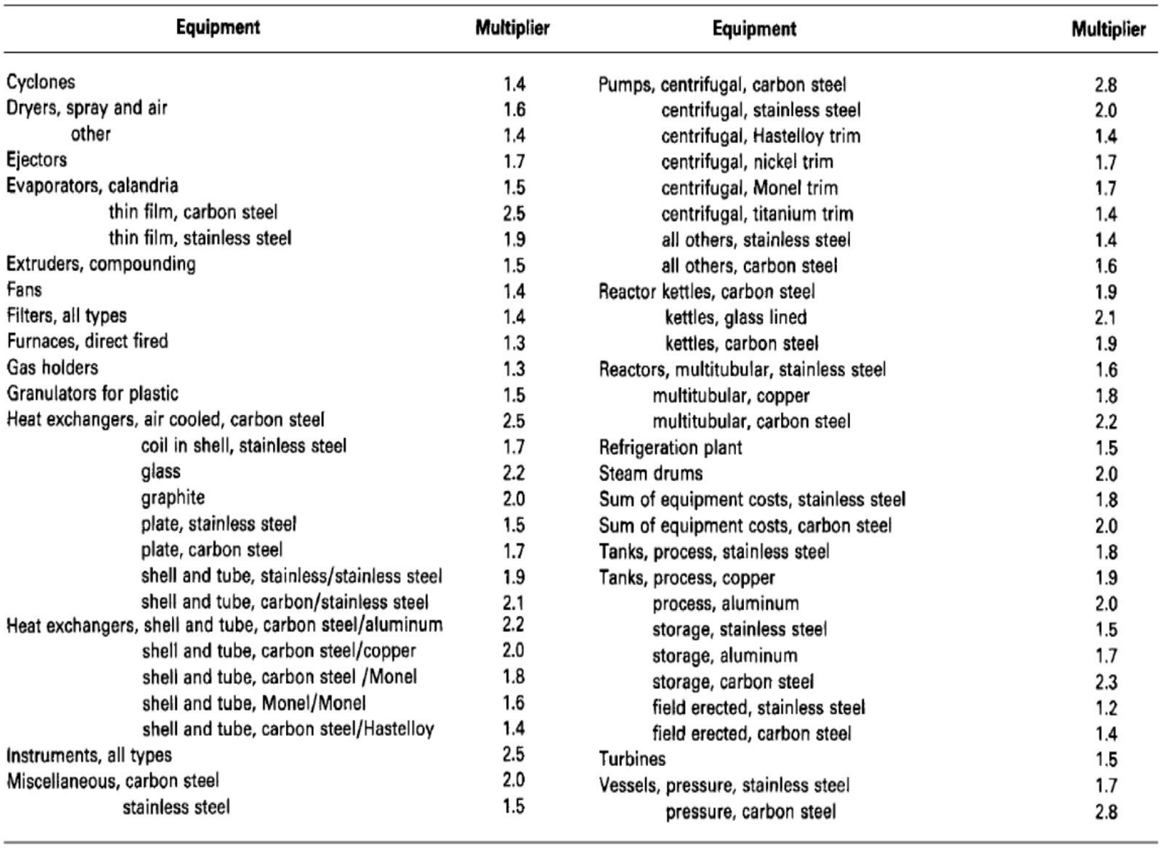
Column Sub-Flowsheet: T-100 @Main (continued)

PROPERTIES

Name	light desiel @Main	heavy desil @Main			
LHV Mass Basis (Std) (kJ/kg)	---	---			
Phase Fraction [Vol. Basis]	0.0000	0.0000			
Phase Fraction [Mass Basis]	0.0000	0.0000			
Partial Pressure of CO2 (kPa)	0.0000	0.0000			
Cost Based on Flow (Cost/s)	0.0000	0.0000			
Act. Gas Flow (ACT_m3/h)	---	---			
Avg. Liq. Density (kgmole/m3)	4.216	3.406			
Specific Heat (kJ/kgmole-C)	491.8	673.7			
Std. Gas Flow (STD_m3/h)	4518	1752			
Std. Ideal Liq. Mass Density (kg/m3)	829.6	859.3			
Act. Liq. Flow (m3/s)	1.459e-002	7.337e-003			
Z Factor	4.704e-003	5.743e-003			
Watson K	11.81	11.79			
User Property	---	---			
Cp/(Cp - R)	1.017	1.012			
Cp/Cv	1.130	1.127			
Heat of Vap. (kJ/kgmole)	8.876e+004	1.059e+005			
Kinematic Viscosity (cSt)	0.5217	0.4527			
Liq. Mass Density (Std. Cond) (kg/m3)	827.9	857.9			
Liq. Vol. Flow (Std. Cond) (m3/h)	45.42	21.80			
Liquid Fraction	1.000	1.000			
Molar Volume (m3/kgmole)	0.2749	0.3564			
Mass Heat of Vap. (kJ/kg)	451.1	419.6			
Phase Fraction [Molar Basis]	0.0000	0.0000			
Surface Tension (dyne/cm)	16.72	15.15			
Thermal Conductivity (W/m-K)	0.1040	0.1003			
Viscosity (cP)	0.3735	0.3205			
Cv (Semi-Ideal) (kJ/kgmole-C)	483.5	665.4			
Mass Cv (Semi-Ideal) (kJ/kg-C)	2.457	2.637			
Cv (kJ/kgmole-C)	435.4	597.8			
Mass Cv (kJ/kg-C)	2.213	2.369			
Cv (Ent. Method) (kJ/kgmole-C)	445.6	613.4			
Mass Cv (Ent. Method) (kJ/kg-C)	2.264	2.431			
Cp/Cv (Ent. Method)	1.104	1.098			
Reid VP at 37.8 C (kPa)	3.877e-002	4.770e-004			
True VP at 37.8 C (kPa)	6.506	6.468			
Liq. Vol. Flow - Sum(Std. Cond) (m3/h)	45.42	21.80			
HHV Molar Basis (Std) (kJ/kgmole)	---	---			
HHV Mass Basis (Std) (kJ/kg)	---	---			
CO2 Loading	---	---			
CO2 Apparent Mole Conc. (kgmole/m3)	---	---			
CO2 Apparent Wt. Conc. (kgmol/kg)	---	---			
Phase Fraction [Act. Vol. Basis]	0.0000	0.0000			
Mass Exergy (kJ/kg)	59.30	109.3			
Partial Pressure of H2S (kPa)	0.0000	0.0000			
Viscosity Index	-7.583	-10.69			

Appendix C

Table C.1 Multipliers for Installed Costs of Process Equipments
(J. Gran , Chem. Eng. , (6 Apr. 1981))



Equipment	Multiplier	Equipment	Multiplier
Cyclones	1.4	Pumps, centrifugal, carbon steel	2.8
Dryers, spray and air	1.6	centrifugal, stainless steel	2.0
other	1.4	centrifugal, Hastelloy trim	1.4
Ejectors	1.7	centrifugal, nickel trim	1.7
Evaporators, calandria	1.5	centrifugal, Monel trim	1.7
thin film, carbon steel	2.5	centrifugal, titanium trim	1.4
thin film, stainless steel	1.9	all others, stainless steel	1.4
Extruders, compounding	1.5	all others, carbon steel	1.6
Fans	1.4	Reactor kettles, carbon steel	1.9
Filters, all types	1.4	kettles, glass lined	2.1
Furnaces, direct fired	1.3	kettles, carbon steel	1.9
Gas holders	1.3	Reactors, multitubular, stainless steel	1.6
Granulators for plastic	1.5	multitubular, copper	1.8
Heat exchangers, air cooled, carbon steel	2.5	multitubular, carbon steel	2.2
coil in shell, stainless steel	1.7	Refrigeration plant	1.5
glass	2.2	Steam drums	2.0
graphite	2.0	Sum of equipment costs, stainless steel	1.8
plate, stainless steel	1.5	Sum of equipment costs, carbon steel	2.0
plate, carbon steel	1.7	Tanks, process, stainless steel	1.8
shell and tube, stainless/stainless steel	1.9	Tanks, process, copper	1.9
shell and tube, carbon/stainless steel	2.1	process, aluminum	2.0
Heat exchangers, shell and tube, carbon steel/aluminum	2.2	storage, stainless steel	1.5
shell and tube, carbon steel/copper	2.0	storage, aluminum	1.7
shell and tube, carbon steel /Monel	1.8	storage, carbon steel	2.3
shell and tube, Monel/Monel	1.6	field erected, stainless steel	1.2
shell and tube, carbon steel/Hastelloy	1.4	field erected, carbon steel	1.4
Instruments, all types	2.5	Turbines	1.5
Miscellaneous, carbon steel	2.0	Vessels, pressure, stainless steel	1.7
stainless steel	1.5	pressure, carbon steel	2.8

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