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Sudan University of Science and Technology

College of Petroleum Engineering and Technology

Department of Transportation and Refining Engineering

Study the effect of variation of crude assay on the design of distillation tower

دراسة تأثير الاختلاف في خواص الخام علي تصميم برج التقطير الجوي

Dissertation submitted in partial fulfillment of the requirement for the Bachelor of Engineering (Horns) Degree in Transportation and Refining Engineering

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Date: / /2014

الاستهلال

قال تعالى:

﴿وما أوتيتم من العلم إلا قليلا﴾

صدق الله العظيم

سورة الإسراء (85)

DEDICATION

We avail this opportunityalso to dedicate this work to our families and specially our parents ,and also to our colleagues in the departments of refining and transportation engineering .

AKNOWLEDGEMENT

First of all , we would like to deeply express our gratitude to supervisor, Dr. Abdelgadir Bashir Banaga for his invaluable guidance , support and we greatly appreciate his cooperation during the stage of writing up this work .

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Abstract

The purpose of this study is to investigate the effect of variation of crude assay on the design of crude distillation column. Different crudes assay (6 samples of Sudanese crude oil) with different properties which are processed at the KRC have been simulated using Aspen Hysys under the same operating conditions in order to identify the effect of this variation . Energy consumption optimization is done by using preheated train heat exchangers to raise the temperature of the crude from 31.2C^0 to 198.3C^0 (124500kw has been saved) . The simulation process show different products quantities with slightly variation in their purities. Specific products (e.g. naphtha) have been increased from 9.1% to 11.8%. As a result of that atmospheric residue decreases from 63.8 % to 60.9% by controlling the reflux ratio , pumps around flow rate and using multi feed locations. The design results are 2.4m column diameter , 27.6 m height of the column and 54 number of trays .

Key word:CDU, Simulation, Design, Control, Hysys

المستخلص

الغرض من هذه المشروع هو دراسة وتقصي تأثير الاختلاف في خواص الخام علي تصميم برج التقطير الجوي ،أجريت عملية المحاكاة لخامات مختلفة المواصفات (6 عينات مختلفة من الخام السوداني) التي تعالج في وحدة التقطير الجوي في مصفاة الخرطوم باستخدام برنامج الهايسس ، مع تثبيت الظروف التشغيلية لمعرفة تأثير هذا الاختلاف ، تم الاستفادة من طاقة المنتجات في تسخين الخام من درجة حرارة 31.2م إلي درجة حرارة 198.3م (تم توفير طاقة مقدارها Kw124500) أوضحت نتائج هذه المحاكاة اختلاف في كمية منتجات برج التقطير الجوي مع اختلاف بسيط في نقاوة هذه المنتجات تم العمل علي زيادة كميات منتجات معينة (نافثا) من 9.1% إلي 11.8% و نتيجة لذلك قلت كمية (المتبقي) من 63.8% إلي 60.9% بالتحكم في عدد من المتغيرات مثل نسبة الراجع ، معدل السريان للمضخات حول البرج و إدخال الخام إلي البرج في عدد من المناطق ، نتائج التصميم هي قطر البرج 2.4م ، ارتفاع البرج 27.6م وعدد 54 صينية .

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Chapter One

Introduction

1.Introduction

1.1 background

Most of Sudanese crude are known for their good quality such as low sulfur content smaller than 0.5wt% and moderate to high an API however like any paraffinic crude oil , some Sudanese crudes has high content of paraffin waxes . Nile blend has an API gravity higher than (>32) , which subsequently declined to 30 API indicating a somewhat heavier crude (medium) , Nile blend has less than 0.06 weight percent of sulfur consider as sweet oil with a high level of wax content (>30%) and with continuous increase in production processes from different regions , some crudes which are considered much heavier than others have contributed in the total production in economical quantities .

AL-FULLA crude which is produced from western kordoffan state can be taken an example of these crudes which has such properties high density and viscosity, high acid value and water content, high calcium content, these properties need to be reduced to the minimum in order to be treated.

Refineries are designed to process a range of crude oils such that their feedstock will provide specific fractions of refined products, sometimes this range will vary greatly form refinery to refinery. The change in crude oil quality around the world has impact the petroleum refining industry in such a way that the current and new refineries are being re-configured and designs respectively to process heavier feedstock, the crudes which are considered heavier than other crudes have to be refined here instead of exporting it as crude to achieve better economic benefits.

1.2 Distillation process

Crude petroleum as it is produced from the field is a relatively low value material since, in its native state, it is rarely usable directly. However, it can be refined and further processed into any number of products whose value is many times that the original oil. The first step in any petroleum refinery is the separation of the crude into various fractions by the process of distillation (physical separation of a mixture into two or more products that have different boiling points). These fractions may be products in their own right or may be feedstock for other refining processing unit.

In most refineries, this process is carried out in two stages. The oil if first heated to maximum temperature allowable for crude being processed and for operation being practiced and then fed to a fractionating tower which operates at slightly above atmospheric pressure. This tower called atmospheric tower. It yields several distillate products and a bottoms product which is the residual liquid material which could not be vaporized under the condition of temperature and pressure existing in the atmospheric tower.

While distillation is one of the important unit operations, it’s one of the most energy extensive operations, the largest consumer of energy in petroleum and petrochemical processing (in some cases distillation is the most economical separation method in liquid mixture however, it could be energy intensive, it can consumed 50% of plant’s operation cost energy) distillation is specialized technology, and the correct design of distillation equipment’s are not always a simple task.

1.3 Distillation towers

Distillation tower in the refinery are designed to process crudes within a certain range of crude characteristics ; crude which are differ in their characteristics to some extend from crudes originally processed in these distillation tower, cannot be processed together , they have been subjected to some treatment processes before processed , or making some designing changes according to the impact , because if they processed together in the same tower it will lowers the distillation efficiency , and increase the overall cost of distillation .

Distillation towers can be classified into two main categories, based on their mode of operation. The two classes are batch distillation and continuous distillation ,In batch distillation, the feed to the column is introduced batch-wise. The column is first charged with a ‘batch’ and then the distillation process is carried out. When the desired task is achieved, the next batch of feed is introduced. Batch distillation is usually preferred in the pharmaceutical industries and for the production of seasonal products. On the other hand, continuous distillation handles a continuous feed stream.

No interruption occurs during the operation of a continuous distillation column unless there is a problem with the column or surrounding unit operations. Continuous columns are capable of handling high throughputs. Besides, additional variations can be utilized in a continuous distillation column, such as multiple feed points and multiple product drawing points. Therefore, continuous columns are the more common of the two modes, especially in the petroleum and chemical industries.

1.4Crude compositions and the effect on design

1.4.1Crude oil assay

Crude oil assay testing includes crude oil characterization of whole crude oils and the boiling – range fractions (TBP distillation test) produced from physical distillation by various procedures. Petroleum assay data are used by clients for detailed refinery engineering and crude oil marketing, feedstock assay data helps refineries optimize the refining process.

1.4.2 The effect on design

crude oil characteristics has become somewhat different from one to another , there is no two crudes oil properties are alike ,so the composition of the total mixture in terms of elementary composition doesn’t vary a great deal , but small difference in composition can greatly affect the physical properties and processing required to produce marketable products .by the time the crude properties may be subjected to a particular change , new discovered wells with different properties contribute the total production , so the overall characteristics could be change by the time , this variation affect the column design and the outcomes of distillation process (change in the products amount and it’s quality) , so some designing change and modification should be done in order to get better separation efficiency.

1.5 project objectives

As the difference in the crude properties increase between some crudes or one crude oil properties change by time the need to design another distillation tower will definitely increase to get more suitable products hence the objective is :

- To study the effect of variation of crude assay in the design of distillation tower.
- Determine the parameters which are use to control the distillation column to deal with variation of crudes assay .
- To Study the optimum choice of treatment to handle with the variation effects ether controlling operating parameters or changing the crude oil properties through blending .

Chapter two

Literature Review

2. Literature review

2.1Reasons for Blending Crude Oils at a Refinery:

Oil refineries, and especially their on-process units that process the primary crude feed stocks, are not uniform in either design or operation. The refinery process units and the combination of units built and in service at a given refinery location are part of a plan to accommodate a certain slate of crudes based on their properties. These decisions are also balanced against the availability and cost of crudes to determine an operating profit point. The more consistent the supply of crude oil to a specific refinery, the more that refinery can tailor its operation to that specific crude supply. However, economics makes that level of optimization difficult to achieve or sustain. Necessity forces refiners to have to retain some flexibility in the refinery process to handle a wider range of crude types than that preferred. Crude blending works hand in hand with refinery process flexibility in crude types by enabling the ability to mix crudes that may not, as individual feeds, satisfy the operating range of the refinery, but as components of a mixed feed will meet the refinery operating requirements.

Refineries can change themselves to accommodate a different crude slate in one of two ways. If the properties of the crude vary only in a minor amount then process cut points, charge rate, and operating set points of existing units can accommodate the changes. Depending on the degree of crude component change, these alterations can be done through the operating conditions or through alterations of the process units during regularly scheduled unit turnarounds. More dramatic changes in crude processing may require the actual construction of different process units at a refinery, an expensive and time consuming step, but one that definitely changes the range of crudes a given site can process.

2.1.1The practical implications of crude blending

The practical implications of crude blending are intermixed with the requirements of purchasing and market analysis. These different, often conflicting requirements, determine the optimal crude slate and refinery operation at any given time, but reflect constantly changing conditions due to :

- Refiners may have based the original design of a refinery on a local oil field characteristic's, only to find after many years of operation that the local field yield has declined, forcing them to source other crudes. The properties of no two crude fields are identical, so no direct replacement is available. However, a blend of two or three available crude oils may come quite close to being the same in properties as the original field. In fact, some crude blending occurs at the field gathering stage itself, as different wells in the same crude field may even have varying properties.
- Even if the local field yield is still capable of supporting the refinery, changing market demands may force a refinery to change its crude source to be able to create more of the range of finished products that are demanded in the local market.
- Refiners would always like to run a crude slate that is low in source sulphur and high in component materials that match local market demand. That is why there is a cost premium for such crudes. By contrast, crude oils high in sediments, sulphur and other contaminants.

- Low in preferred component materials are sold at discount. Much of the more recent oil fields and supplies coming online have been of these heavy crudes. For a refinery that can't process the heavy crudes, costs are driven up by premium to purchase light-sweet crudes. However, a refinery that can't run pure heavy crude may still be able to run some heavy crude when it is diluted into the light-sweet crude. This allows the refinery to purchase some percentage of their crude at lower prices.
- At current refinery operating margins, nothing can be wasted. Off specifications and slops materials are reprocessed when they are not capable of being sold as product, or reprocessed when the available prices of those products aren't viable. However, no crude process unit can afford to receive a

feed stream of undiluted slops and seconds, the variability of the properties is too high.

- International politics can play havoc with the supply of crudes. Refiners have been forced in the past to change their crude sourcing on a moment’s notice for no economic reason, and the odds are will have to again. Changing the source region or country for the origin of crude is a similar problem to the original field or crude slate type becoming economically unviable, the refiner is forced to source what is available at competitive prices, and figure out a mix of those that corresponds to the refinery process capabilities.
- Crude blending may even be done by terminals and suppliers prior to a refinery. Under this scenario, crude blending is done to meet a refiner’s target spec range for crude. Question arise about how to best co-ordinate plant optimization with crude slate optimization when separate entities are involved .(Kevin,2004)

2.2 Previous research

2.2 .1 Future technology in heavy oil processing

The change in crude oil quality around the world has impact the petroleum refining industry in such a way that the current and new refineries are being re-configured and design respectively to process heavier feedstock .blends of various crude oil with elevated amount of heavy petroleum this is due to the reduction of light crude oil and increase of heavy or extra crude oils production .these new feeds are characterized by high amount of impurities , low distillate yields , which make them more difficult to process compared with light crude oils, contrarily, the demand for light distillate for producing the so called clean fuels is increasing throughout the world .these circumstances situate not only refineries but also research need to adapt and design future technologies for properly conversion and upgrading of heavy oils there are various available technologies which can be classified in carbon rejection , hydrogen addition process .(**Jorge , 2005)**

2.2.2 Crude oil blending

Refineries are designed to process a range of crude oils such that their feedstock will provide specific fractions of refined products. Sometimes this range will vary greatly from refinery to refinery. The various processing units are designed and sized so that all the running near capacity for a given feedstock stream. The plant’s efficiency is maximized in doing so. If the feedstock stream changes then therein the cut or composition of the various processed streams. An example would be a light oil refinery that has asphalt-making capabilities and therefore would require a heavy component in its feedstock. If the regular feedstock stream changed so that it didn’t have this heavy component then a heavy stream would need to be added into the main feedstock Blending in this manner allows the refiner to match the incoming feedstock with existing equipment **. (Kevin , 2004).**

2.2.3Manufacturing Light Oil from Heavy Crude Ratqa Field, North Kuwait

Heavy Oil from North Kuwait does not have an intrinsic commercial value by itself. The crude is estimated to have an API in the 11-18 API range and high sulphur of >5% wt., which makes extremely difficult the processing operation in a conventional crude oil refinery. Notwithstanding, currently there are two options to make this crude marketable the first is by diluting or Blending the Heavy Oil (11 API) with a much lighter crude oil to produce a blend to be placed in the open market , the second by processing the Heavy Oil in Upgrading complexes and, depending on the selected upgrading scheme, produce a range of upgraded crude oils which vary in API from 16 to 35 and have sulphur content between 0.1 and 3.5 % wt, The technical challenge for studying this option (Upgrading an 11-15 API crude to 31 API) is to find the optimal upgrading technology and to integrate an optimal upgrading scheme capable of manufacturing an upgraded crude oil with a commercial market value, as one option to dispose the Heavy Oil. That upgrading scheme is presented, in this paper, as an option for disposition of the HO from North Kuwait**.(Luis , 2009).**

2.2.4 Continues distillation column

In the control of batch and continuous distillation columns, one of the most challenging problems is the difficulty in measuring compositions. This problem can be handled by estimating the compositions from readily available online temperature measurements using a state observer. The aim of this study is to design a state observer that estimates the product composition in a multicomponent batch distillation column (MBDC) from the temperature measurements and to test this observer using a batch column simulation. The effect of measurement period value is also studied and found that it has a major effect on the performance which has to be determined by the available computational facilities. The control of the column is done by utilizing the designed EKF estimator and the estimator is successfully used in controlling the product purities in MBDC under variable reflux-ratio operation (**Yıldız, 2002**).

2.2.5 Operation and control high purity distillation column

A dynamic model for continuous minimum reflux distillation column is developed for the case of constant relative volatility (CRV) and constant molar overflow (CMO). This model predicts the dynamic behavior of minimum reflux distillation columns and provides a clear description of both the steady state and the dynamic conditions necessary for maintaining high purity operation. Boundaries at which the steady state composition profile in a binary minimum reflux column is indeterminate are identified and analytical expressions predicting the location of these boundaries are established. Two new model based control strategies are presented which provide excellent disturbance rejection for large variations in feed composition. The optimum operating policy for a thermallycoupled ternary (**John, 1996**).

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Another scientistspresented a similar simplified model for batchdistillation, also based on the FUG shortcut method for continuous distillation design. Their model was run at constant overhead composition and constant reflux ratio and was shown to be in excellent agreement with rigorous simulations under the assumption of constant molar overflow and zero liquid and vapour holdup. The main difference between the models of Diwekar and Sundaram is in their input data. The model input data required makes Diwekar's model suited to design while Sundaram's model is applicable also to rating studies on an existing column**(Sundaram and Evans (l993**)).

Several soft-wares have been developed for the petroleum industries. ASPENTM HYSYS is one of the software which is widely accepted and used for refinery simulation. ASPENTM HYSYS performs the oil distillation calculation through detail plate by plate calculation. This work aims to study the quality of three products of a fractionation column 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. (**A. Rahmanajune ,2011).**

This paper aim to investigate the improvements which could be applied to the solution of multicomponent multistage batch distillation problems as a means to get more precise and flexible results from rigorous simulations. Then they employed the simulated responses to verify the numerical solution algorithm, by comparing both (i) with the simulation using Runge-Kutta method and (ii) with the experiment. Their algorithm incorporated a generalized Implicit Euler integration supported with an overall normalized £ method accelerating convergence. It should be noted that, due to assumptions in the base models, the simulation starts from the total-reflux steady-state, which prevents the simulation from giving the dynamic behavior during the start-up period. zThe algorithm resulted in a higher accuracy and lesser CPU-time than Runge-Kutta integration , as well as comparing the simulation results with that from a rigorous model, also compared it with experimental data. In many cases, the shortcut model was seen to compare quite well with the experimental and rigorous models and thus lead to savings in computational effect **(Mori *et al.* (1995))**.

Chapter Three

Methodology

3.Methodology

3.1 Case study

Different Sudanese crude oils with different assay from different fields mixed crude oil of HEGLIG, unity, andthargas has being processed in one atmospheric distillation column in order to realized the effect of this variation on the design process , by fixing the same operating conditions of all the units that the crude pass through which are the preheated trains , pr-flash separator , furnace and the two side strippers (steamed stripped) for the LGO and HGO in addition to one side stripper of kerosene (reboiled striped) . All crudes has been simulated using HYSYS as a simulator tool.

Main properties of the 6 samples which are processed at KRC has been simulated on hysis in our project :

Table 3.1The Nile Blended Crude Oil sample 2002 **(sample 1)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	0.583
API Gravity	ASTM D5002	-	34.3
Total Sulphur	ASTM D4294	% wt	0.05

Table 3.2The Nile Blended Crude Oil sample 2004**(sample2)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	0.855
API Gravity	ASTM D5002	-	33.91
Total Sulphur	ASTM D4294	% wt	0.056

Table 3.3The Nile Blended Crude Oil sample **2006(sample3)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	0.8504
API Gravity	ASTM D5002	-	34.81
Total Sulphur	ASTM D4294	% wt	0.051

Table 3.4The Nile Blended Crude Oil sample 2010 **(sample4)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	.871
API Gravity	ASTM D5002	-	30.81
Total Sulphur	ASTM D4294	% wt	0.0602

Table 3.5 The Nile Blend and LCO blending Crude Oil sample 2012 (**sample5)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	0.8686
API Gravity	ASTM D5002	-	31.2
Total Sulphur	ASTM D4294	% wt	0.063

Table 3.6 Assay of 90/10 Nile/TharJath Blend (**sample6)**

Test	Method	Unit	Results
Density at 15°C	ASTM D5002	g/cm³	0.8648
API Gravity	ASTM D5002	-	32.04
Total Sulphur	ASTM D4294	% wt	0.073

3.1.1 process description

Fresh feed are preheated using heat exchangers to raise the temperature of the crude from $31.2C^0$ to $198.3C^0$, then the heated crude enter a pre flash tower to separate generated vapor to avoid the explosions inside the furnace after that the crude oil is heated again until his temperature reach $360C^0$ and directly to the atmospheric distillation column and it is separated into different products , Distillate productsare removed from selected trays (draw-off trays) in this sections of the tower. Thesestreams 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 portion of this stream is returned as reflux while the remainder is delivered to the light end process 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)

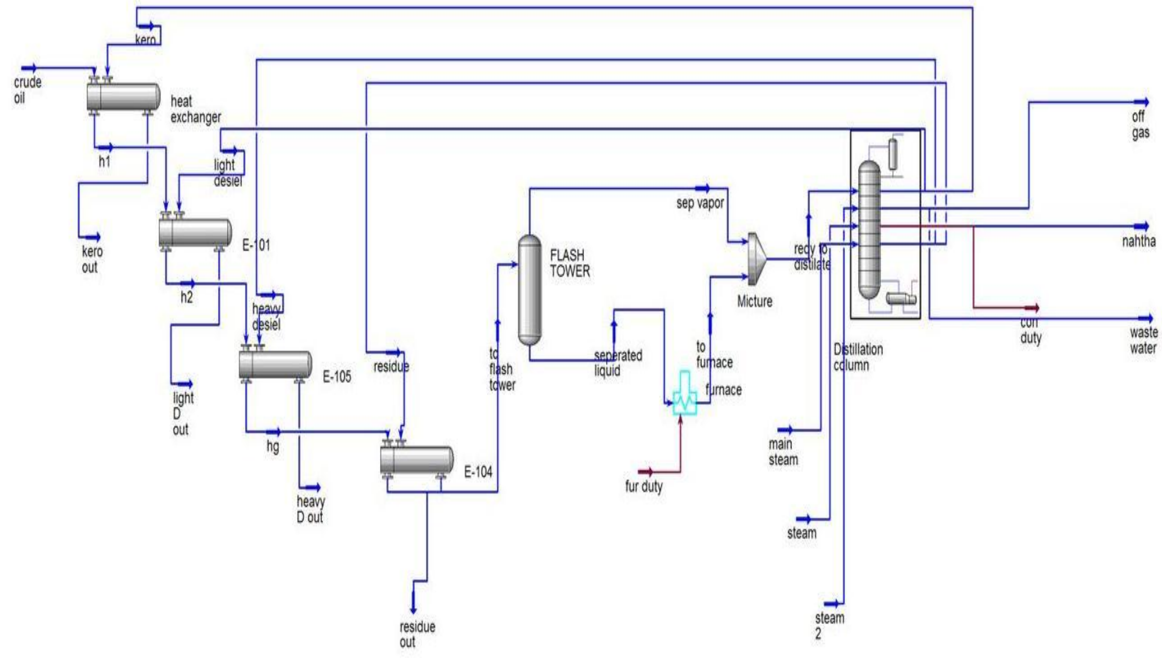


Figure 3.1 process description of crude distillation column .

3.2 Hysys process simulator

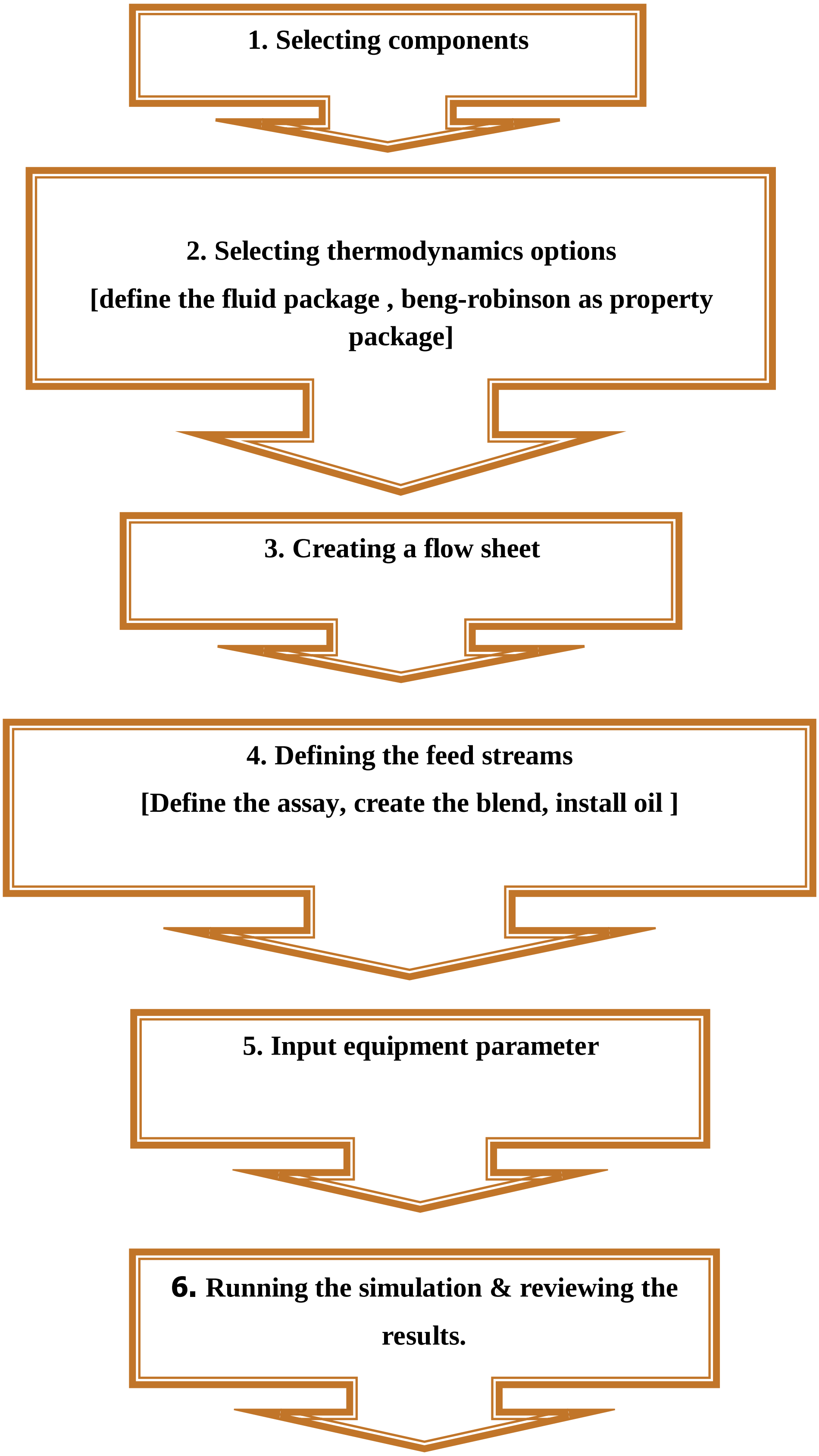
HYSYS is a process simulation environment designed to serve many processing industries especially oil and gas refining. Rigorous steady state and dynamic models for plant design, performance monitoring, troubleshooting, operational improvement, business planning and asset management can be created using HYSYS. The built-in property packages in HYSYS provide accurate thermodynamic, physical and transport property predictions for hydrocarbon,nonhydrocarbon, petrochemical and chemical fluids.

The calculation method for distillation in HYSYS is done to a high standard in accordance with the matrix method. A quick convergence and short simulation time is therefore guaranteed. In most cases, the user need not be concerned with the details of the internal calculation this is done automatically by HYSYS.

3.2.1 Process simulation procedure

The following six basic steps are used to run a flow sheet simulation in HYSYS simulator all these steps are followed to process the different Sudanese crude in order to know the effect of variation of these crude on the distillation column design.

On the six samples energy consumption optimization has been applied to the column by using the hot products from the strippers to raise the temperature of the crude oil before entering the furnace to reduce the energy needed to heat the crude until 198 c using the product and also maintain the temperature of the products on the range below to their flash points .


1. Selecting components
2. Selecting thermodynamics options
[define the fluid package , beng-robinson as property package]
3. Creating a flow sheet
4. Defining the feed streams
[Define the assay, create the blend, install oil]
5. Input equipment parameter
6. Running the simulation & reviewing the results.

3.3 Stepwise Distillation design procedure

3.3.1 Multi-component distillation tower

The problem of determining the stage and reflux requirements for multi component distillations is much more complex than for binary mixtures. With a multi-component mixture, fixing one component composition does not uniquely determine the other component compositions and the stage temperature. Also when the feed contains more than two components it is not possible to specify the complete composition of the top and bottom products independently. The separation between the top and bottom products is specified by setting limits on two "key" components, between which it is desired to make the separation. (R.K.Sinnott,2005)

3.3.2 Key components

Before commencing the column design, the designer must select the two "key"components between which it is desired to make the separation. The light key will be the component that it is desired to keep out of the bottom product, and the heavy key the component to be kept out of the top product Specifications will be set on the maximum concentrations of the keys in the top and bottom products.

The components that have their distillate and bottoms fractional recoveries specified are called key components. The most volatile of the keys is called the light key (LK) and the least volatile is called the heavy key (HK). The other components are called non-keys (NK).

3.3.3 Determine the stage and reflux requirement

3.3.3.1Calculation of Minimum number of stages N_{min}

Fenske (1932) was the first to derive an Equation to calculate minimum number of trays for multicomponent distillation at total reflux. The derivation was based on the assumptions that the stages are equilibrium stages.

Fenske Equation can be easily derived for multi-component calculations which can be written as:

$$N_{min} = \frac{log\,(\frac{xD\,,l}{xD\,,H}\times\frac{xB\,,H}{xB\,,l})}{\log\,(\alpha l\,,H)} \quad \text{---}\text{Eq(3.1)}$$

Where i=LK and j=HK, the minimum number of equilibrium plots is influenced by the components only by their effect on the value of the relative volatility between the key components.

Thus, the minimum number of equilibrium stages depends on the degree of separation of

The two key components and their relative volatility, but is independent of feed-condition where the mean relative volatility is approximated by

$$\left(\alpha l, H = \frac{\alpha ,lk}{\alpha ,Hk} \right)$$

3.3.3.2Calculation of Minimum Reflux Ratio R_m

3.3.3.2.1 Minimum Reflux

Underwood Equations For multi-component systems, if one or more of the components appear in only one of the products, there occur separate pinch points in both the stripping and rectifying sections. In this case, Underwood developed an alternative analysis to find the minimum reflux ratio (Wankat, 1988),Minimum reflux is based on the specifications for the degree of separation between two key components. The minimum reflux is finite and feed, product with drawls are permitted.

However, a column cannot operate under this condition because of the requirement of infinite stages. But it is useful limiting condition.

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q \quad \text{--Eq(3.2)}$$

q = condition of feed

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = Rm + 1 \quad \text{-- Eq(3.3)}$$

3.3.3.3 Calculation of Actual Reflux Ratio

R= (1.1-1.5)*R_{min} --Eq(3.4)

3.3.3.4Theoretical number of stages

Gilliland (1940) developed an empirical correlation to relate the number of stages N at a finite reflux ratio L/D to the minimum number of stages and to the minimum reflux ratio .

Using correlation

$$X = \frac{R-Rmin}{R+1}$$ ---Eq(3.5)

$$Y = (1 - X^{1/3})$$ -------------------------------------Eq(3.6)

Y=

$$1-\exp\left[\left(\frac{1+54.4(R-R_{min})/(R+1)}{11+117.2(R-R_{min})/(R+1)}\right)\left(\frac{(R-R_{min})/(R+1)-1}{[(R-R_{min})/(R+1)]^{0.5}}\right)\right]$$ --------------------------Eq(3.7)

$$N = (N_{min} + Y)/(1-Y)$$ -----------------------------------Eq(3.8)

Using erbar-maddox correlation chart

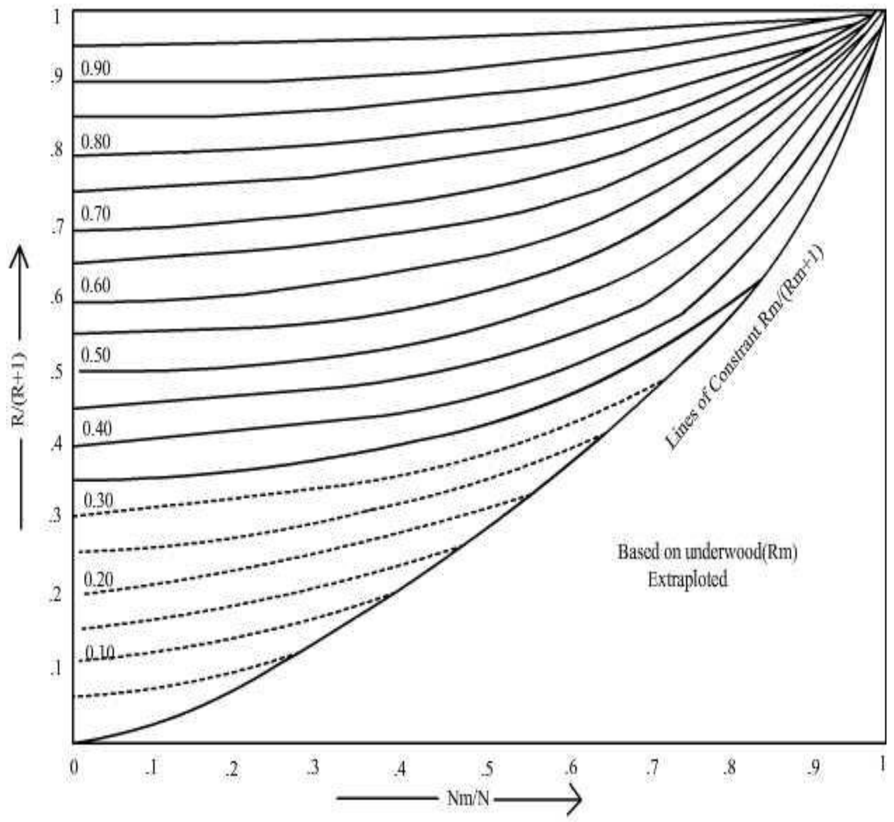


Figure 3.2Erbar-Maddox colleration

3.3.3.5Calculation of actual number of stages

3.3.3.5.1Estimating Efficiencies – The O'Connell Method

There are many methods that have been developed to estimate distillation efficiencies. Here we consider just one method; that of H.E. O'Connell (*Trans. AIChE*, **42**, 741, 1946). O'Connell obtained his correlation for the efficiency of distillation processes from an analysis of data on several operating columns. The original correlation was graphical, but equations have been proposed to represent the correlation. One such equation is

$$E_{oc} = 50.3(\alpha\mu)^{-0.226} \quad \text{------------------------------} \quad \text{Eq(3.9)}$$

Where (α) is the relative volatility between the key components and(μ) is the viscosity in cP. The correlation is shown in the chart below.

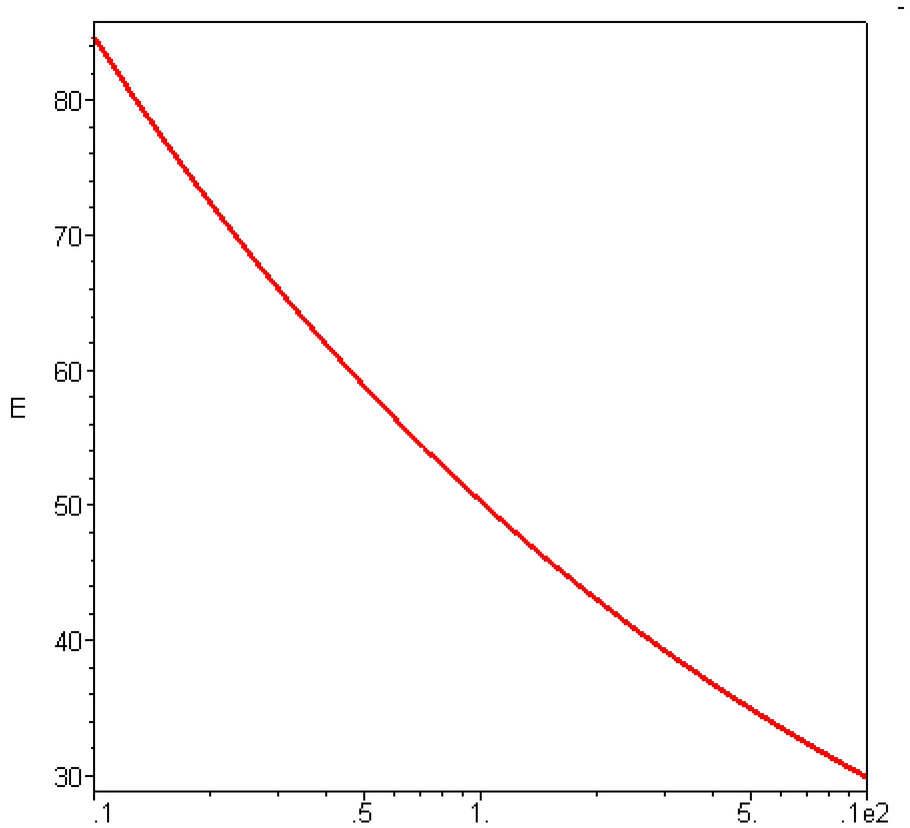


Figure 3.3 relative volatility times viscosity (cp)

Actual number of stages = $\frac{\text{N theoritical}}{\text{E}}$ ----------------------Eq(3.10)

3.3.4Feed plate location

A limitation of the Erbar-Maddox, and similar empirical methods, is that they do not give the feed-point location. An estimate can be made by using the Fenske equation to calculate the number of stages in the rectifying and stripping section sseparately, but this requires an estimate of the feed-point temperature ,an alternative approach is to use the empirical equation given by Kirkbride(1994).

$$\log\left(\frac{N_R}{N_S}\right) = 0.206 \log\left[\left(\frac{D}{W}\right)\left(\frac{x_{f,HK}}{x_{f,LK}}\right)\left(\frac{x_{b,LK}}{x_{d,HK}}\right)^2\right]$$ ----------------------Eq(3.11)

Where:

N_r=number of stages above the feed, including any partial condenser.
N_s= number of stages below the feed, including the reboiler.
B = molar flow bottom product.
D = molar flow top product.
$X_{f,HK}$= concentration of the heavy key in the feed.
$X_{f,LK}$ =concentration of the light key in the feed.
$X_{d,HK}$= concentration of the heavy key in the top product,
$X_{b,HK}$= concentration of the light key if in the bottom product.

3.3.5 Plate spacing

The overall height of the column will depend on the plate spacing. Plate spacing from0.15 m (6 in.) to 1 m (36 in.) are normally used. The spacing chosen will depend onthe column diameter and operating conditions. Close spacing is used with small-diameterColumns, and where head room is restricted; as it will be when a column is installed in abuilding. For columns above 1 m diameter, plate spicing of 0.3 to 0.6 m will normallybe used, and 0.5 m (18 in.) can be taken as an initial estimate. This would be revised, asNecessary, when the detailed plate design is made.

3.3.6Column daimeter

The principal factor that determines the column diameter is the vapor flow-rate. Thevapor velocity must be below that which would cause excessive liquid entrainment or ahigh-pressure drop. The equation given below, which is based on the well-known Souders and Brown equation, Lowenstein (1961), can be used to estimate the maximum allowablesuperficialvapour velocity, and hence the column area and diameter

$$\hat{u}_v = (-0.171 l_t^2 + 0.27 l_t - 0.047)\left[\frac{(\rho_L - \rho_v)}{\rho_v}\right]^{1/2}$$

--------------Eq(3.12)

Where:
$\hat{u}_v$= maximum allowable vapour velocity, based on the gross (total) column cross-sectional area, m/s.

The column diameter Dc, can then be calculated

$$D_c = \sqrt{\frac{4\hat{V}_w}{\pi \rho_v \hat{u}_v}}$$

-----------------------------------Eq(3.13)

Where:
$\hat{V}_w$ =is the maximum vapor rate,m/s.
ρ_v =Vapor density kg/m^3.
D_c = column diameter ,m.

3.3.7 Total pressure drop over the column

The pressure drop over the plates is an important design consideration. There are two main sources of pressure loss: that due to vapour flow through the holes (an orifice loss) and that due to the static head of liquid on the plate.

A simple additive model is normally used to predict the total pressure drop. The total is taken as the sum of the pressure drop calculated for the flow of vapour through the dry plate (the dry plate drop hd); the head of clear liquid on the plate (hw+ how); and a term to account for other, minor, sources of pressure loss, the so-called residual loss hr. The residual loss is the difference between the observed experimental pressure dropand the simple sum of the dry-plate drop and the clear-liquid height. It accounts for the two effects: the energy to form the vapour bubbles and the fact that on an operating plate the liquid head will not be clear liquid but a head of “aerated” liquid froth, and the froth density and height will be different from that of the clear liquid.

3.3.7.1 Tray hydraulic parameters

3.3.7.1.2 Dry plate drop hd

Dry plate pressure drop occurs due to friction within dry short holes$h_d$can be calculated using following expression derived for flow through orifices.

$$h_d = 51 \left[\frac{u_h}{C_0}\right]^2 \frac{\rho_v}{\rho_L} \text{-------------------------------}\text{Eq(3.14)}$$

u_h= vapor velocity through holes m/s

$$u_h = \frac{qv}{Ah} \text{-------------------------------}\text{Eq(3.15)}$$

3.3.7.1.3Residual pressure drop

The residual pressure drop results mainly from the surface tension as the gas releases from a perforation. The following simple equation can be used to estimate $h_r$with reasonable accuracy

$$h_r = \frac{12.5*10^3}{\rho L} \text{-----------------------------}\ \text{Eq(3.16)}$$

Total pressure drop per plate is then cumulatively found from:

$$h_t(m) = h_d + (h_w + h_{ow}) + h_r \text{----------------------------}\ \text{Eq(3.17)}$$

3.4 Effect of vapor flow conditions on tray design

3.4.1 Flooding check

Excessive liquid buildup inside the column leads to column flooding condition. The nature of flooding depends on the column operating pressure and the liquid to vapor flow ratio. The column diameter must be selected so that flooding does not occur , however at the same time one needs vapor velocityies that are higher for great plate efficiencies.

$$u_f = K1\sqrt{\frac{\rho_L-\rho_V}{\rho_L}}\text{----------------------------------}\ \text{Eq(3.18)}$$

u_f flooding vapour velocity, m/s, based on the net column cross-sectional areaA_n.

K1 a constant obtained from Figure

$u_f > U_n$

No flooding occur

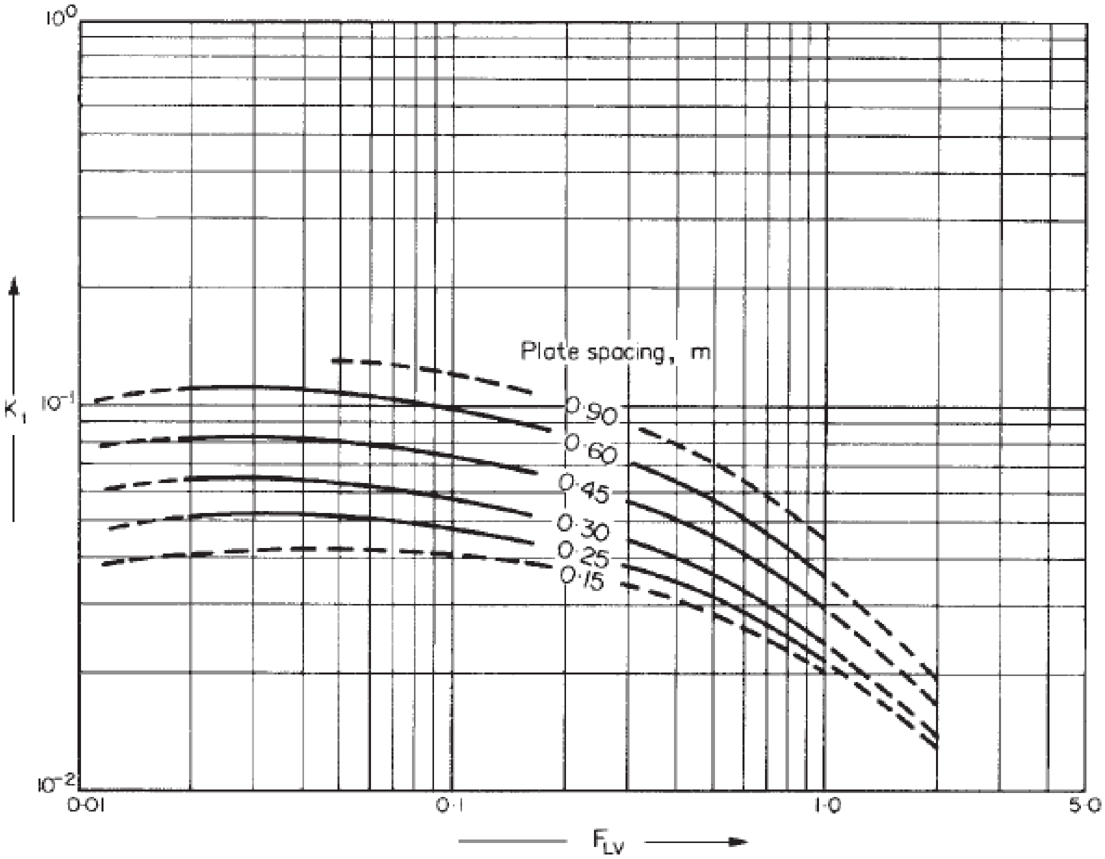
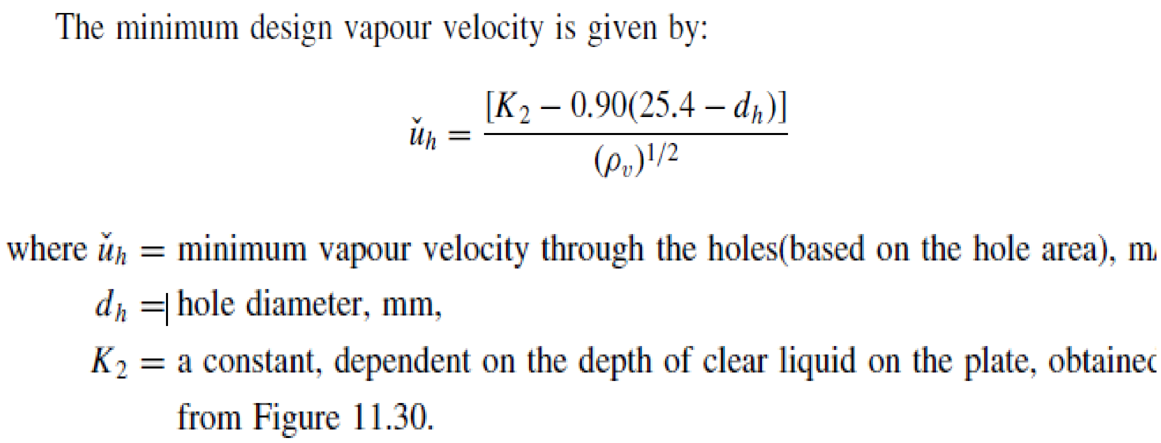


Figure 3.4 fooling velocity, sieve plates

3.4.2 Check of weeping

The lower limit of the operating range occurs when liquid leakage through the plateholes becomes excessive. This is known as the weep point. The vapour velocity at theweep point is the minimum value for stable operation.



The minimum design vapour velocity is given by:

$$\breve{u}_h = \frac{[K_2 - 0.90(25.4 - d_h)]}{(\rho_v)^{1/2}}$$

where $\breve{u}_h$ = minimum vapour velocity through the holes(based on the hole area), m

d_h =| hole diameter, mm,

K_2 = a constant, dependent on the depth of clear liquid on the plate, obtaine from Figure 11.30.

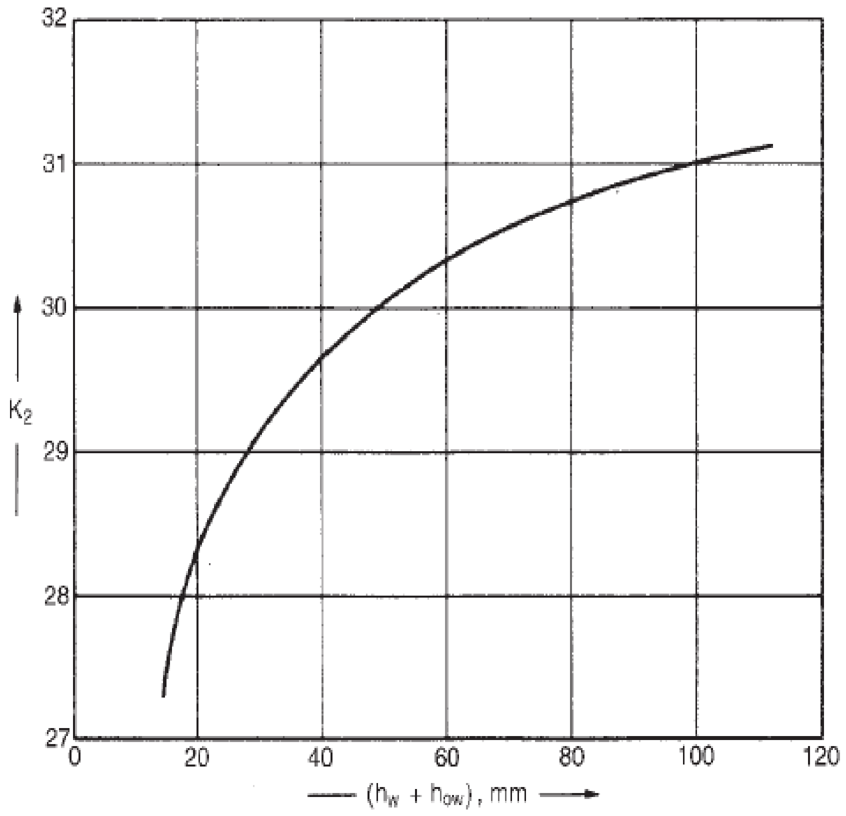


Figure 3.5 weep point correlation

$$U_{\text{min ,op}} = \frac{minimum \ \ vapor \ flow \ rate}{hole \ area}$$ -----------------------Eq (3.19)

To avoid weeping$U_{min\ .op} > U_{min}$

Weeping will not take place.

3.4.3 Check of fractional entrainment

Entrainment is the phenomena in which liquid droplets are carried by vapor/gas to the tray above Entrainment can be estimated from the correlation given by Fair (1961), Figure 11.29, which gives the fractional entrainment ψ (kg/kg gross liquid flow) as function of the liquid-vapour factor FLV, with the percentage approach to flooding as a parameter. The percentage flooding is given by:

Percent flooding = U_n/ U_f

ifψ ≤0.2 the design is ok

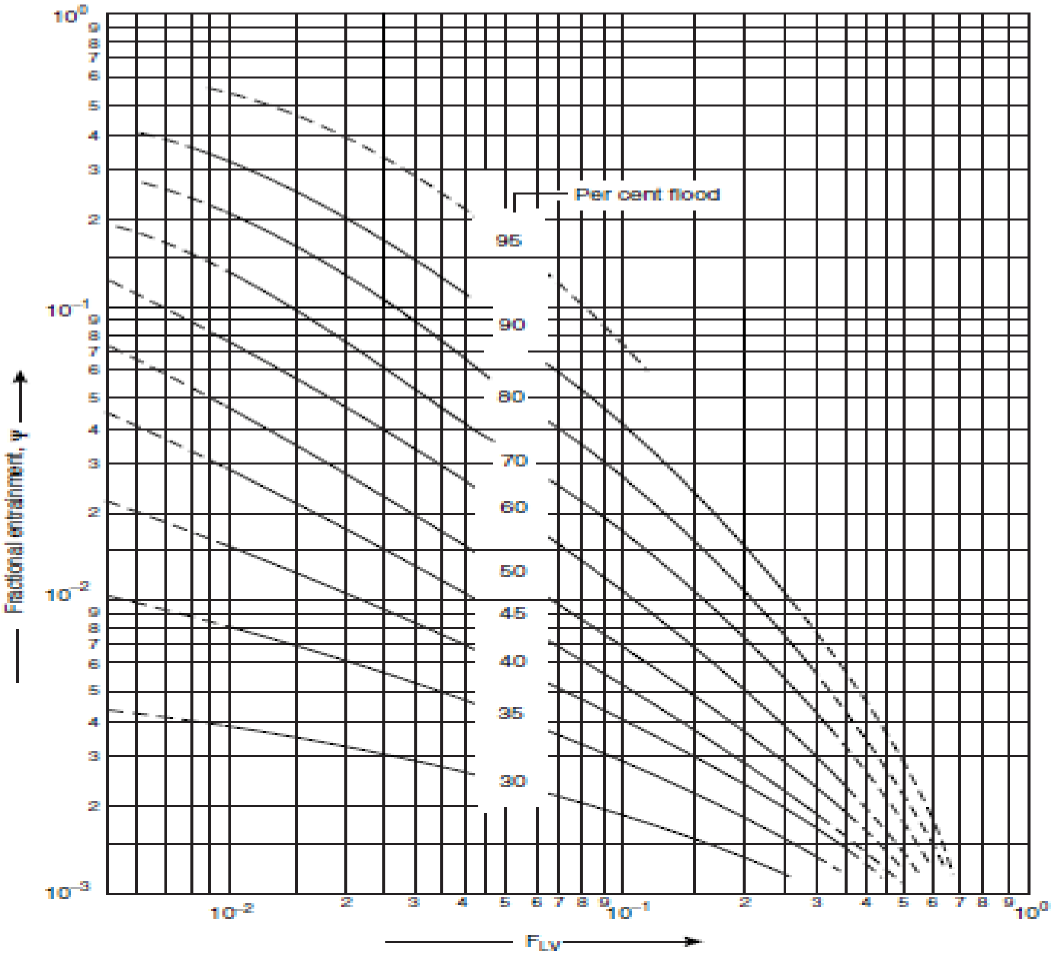


Figure 3.6 Entrainment correlation for sieve plate

3.4.4 Check of residence time in the downcomer

Sufficient residence time must be allowed in the downcomer for the entrained vapor toDisengage from the liquid stream; to prevent heavily “aerated” liquid being carried underthe downcomer.

L_d= H_w+C

If Ţ>3s , ok

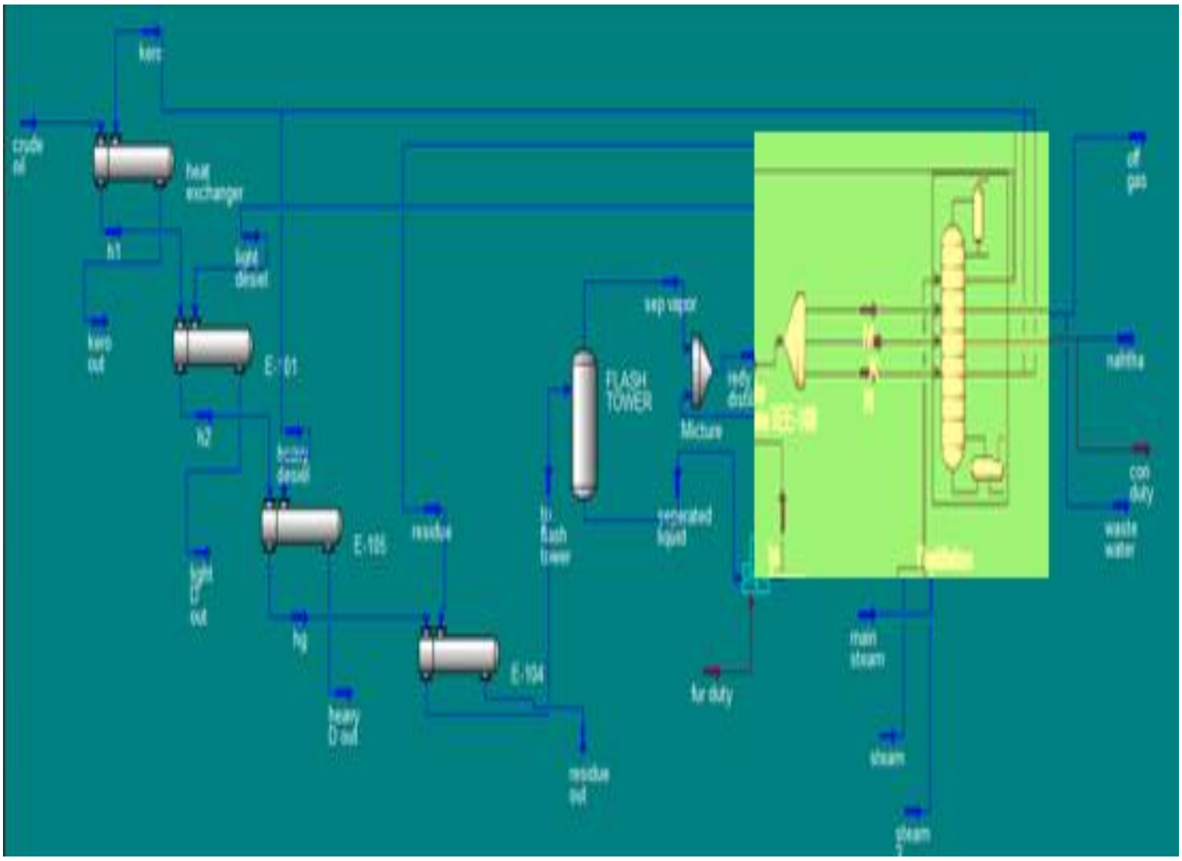
3.5 Methodology of studying the effect of increasing the reflux ratio

Form the program monitor changing the reflux ratio fixing other operating condition Constant.

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	✓	✓

3.6Methodology of using multi feed positions

Instead of entering the feed on the third plate (one location) the feed has been divided (30%,30%,40%) and enter to the tower at different location (8,5,3respectively) fixing the other parameters constant


kero
crude oil
heat exchanger
h1
light diesel
kero out
E-101
h2
heavy diesel
E-105
light D out
residue
hg
heavy D out
E-104
residue out
to flash tower
FLASH TOWER
sep vapor
redy dist
Micture
separated liquid
fur duty
off gas
naphtha
con duty
waste water
main steam
steam
steam 2
Distillation

Chapter Four

Results & Discussion

4. Results and discussion

Introduction

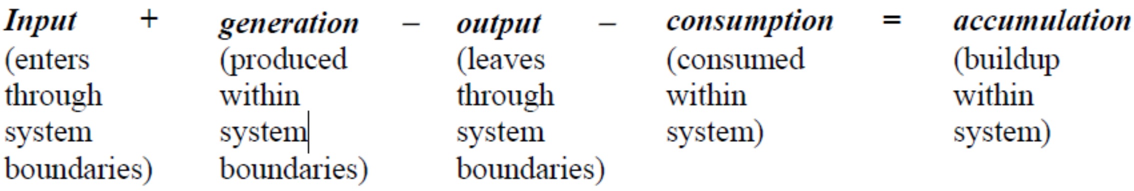
In order to identify the effect of variation of crudes assay on the design of distillation column there are some significant calculation related to the design of distillation column and the auxiliary equipment in the CDU.These calculations include: calculations of material and energy balance , multi component distillation column design calculations and cost estimation calculations.

4.1calculation of Material balance

Material balances are important first step when designing a new process or analyzing an existing one. They are almost always prerequisite to all other calculations in the solution of process engineering problems.Material balances are the basis of process design. A material balance taken over the complete process will determine the quantities of raw materials required and products produced. Balances over individual process units set the process stream flows and compositions, Material balances are nothing more than the application of the law of conservation of mass, which states that mass can neither be created nor destroyed

4.1.1The General Balance Equation

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



Input	+	*generation*	–	*output*	–	*consumption*	=	*accumulation*
(enters through system boundaries)		(produced within system boundaries)		(leaves through system boundaries)		(consumed within system)		(buildup within system)

The general balance equation may be simplified according to the process at hand. For example, by definition, the accumulation term for steady-state continuous process is zero thus the above equation becomes:

Input + generation = output + consumption

For physical process, since there is no chemical reaction, the generation and consumption terms will become zero, and the balance equation for steady-state physical process will be simply reduced to*:*

Input = Output

A CDU produces five different products namely [gas + naphtha (GN), Kerosene (K), Light gas oil (L), heavy gas oil (H) and residue (R)] .

The steady volumetric balance for the CDU is defined as:

$F_{crude} = F_{NG} + F_K + F_L + F_H + F_R$ -------------------------Eq (4.1)

Where F refers to the volumetric flow rates of various streams (crude, GN, K, L, H and R).

The mass balance for the CDU is defined as:

$MF_{crude} = MF_{NG} + MF_K + MF_L + MF_H + MF_R$

Where MF refers to the mass flow rates associated to the feed and product streams.

The mass balance in this chapter, it's consist of:

- Balance around flash tower.
- Balance around distillation.

4.1.2Material balance around Flash tower

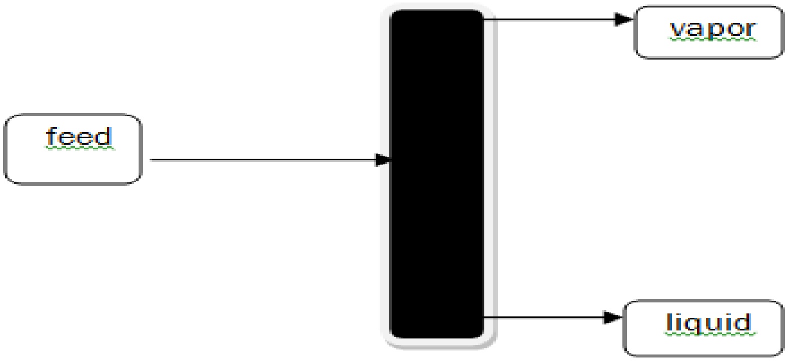


figure 4.1 flash tower

M=feed

V=vapor flash tower

L=liquid flash tower

Total feed =258 t/h

V/M=4.202/258 =0.0162868

Vapor flash tower = 258000*0.0162868=4202 kg/h = 4.202 t/h

Liquid flash tower = 258-4.202=253.798 t/h

4.1.3 Material balance around distillation column

The equation Material balance:

Input +generation – consumption –output =accumulation

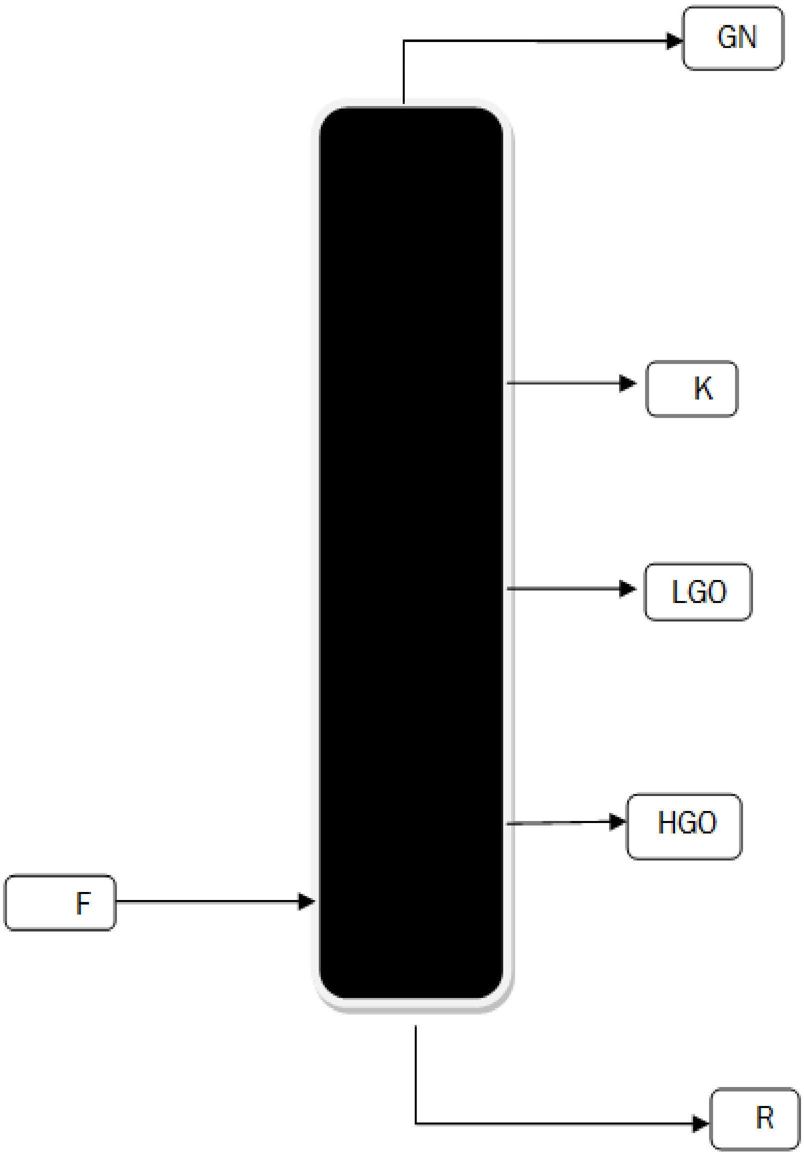


Figure 4.2 distillation tower

F=feed

GN= gas + naphtha

K= kerosene

LGO= light gas oil

HGO= heavy gas oil

R=residue

Calculation of product rat in ton/h

The mass flow rate of (NG)

NG=258*0.07945=20.5 ton/h

The mass flow rate of kerosene

K=258*0.03877=10 ton/hr

The mass flow rate of LGO

LGO=258*0.14573=37.6 ton/hr

The mass flow rate of HGO

HGO=258*0.07248=18.7 ton/hr

The mass flow rate of Residue

R=258*0.66356= 171.2 ton/h

Table 4.1Overall material balance of distillation tower

Properties	**Unit**	**Whole Crude@ 15 C**	**N**	**K**	**LGO**	**HGO**	**R**
% in feed	%	100	7.945	3.877	14.574	7.248	66.356
Mass flow (MF)	(t/h)	258	20.5	10	37.6	18.7	171.2
Mass flow rate	kg/s	71.67	5.694	2.778	10.44	5.194	47.555
Density	kg/m^3	855	675.6	749	757	770.2	849.2
Specific gravity	--	0.855	0.6756	0.749	0.757	0.7702	0.8492
Volumetric flow rate	m^3/s	0.0838	0.00843	0.00304	0.0138	0.00675	0.0559
Molecular weight	--	261.147	92.7355	166.435	237.1	282.37	413.226
Molar flow	kmol/h	987.94	253.17	139.513	143.5	46.59	395.317

4.1.4calculation of Overall Water material balance:

At steady state

Water in = water out

Table 4.2 Calculation of Overall Water material balance

	In				Out						
Stream	Crude	St1	St2	Main st	Off gas	Naphtha	Waste water	kero	Light Diesel	Heavy Diesel	Residue
Percent %	0	1	1	1	6.17	0.0493	1	0	0.264	2.918	0.239
Amount in (kmol/h)	0	127.3	71.61	127.3	0.944	0.13620	0.379	0	265.7	0.125044	0.7450
Amount in (kg/h)	0	1290	129	2294	17.02	2.45443	6.839	0	4787.914	2.253287	13.426
sumation	Nput	I 270.5 (kmol/) 4.8 (ton/h)				Out put		270.5 (kmol/) 4.8 (ton/h)			

4.2 Calculation of Energy balance

4.2.1 Introduction

The conservation of energy differs from that of mass in that energy can be generated (or consumed) in a chemical process.

Material can change form, new molecular species can be formed by chemical reaction, but the total mass flow into a process unit must be equal to the flow out at the steady state. The same is not true of energy. The total enthalpy of the outlet streams will not equal that of the inlet streams if energy is generated or consumed in the processes; such as that due to heat of reaction.

Energy can exist in several forms: heat, mechanical energy, electrical energy, and it is the total energy that is conserved.

In process design, energy balances are made to determine the energy requirements of the process: the heating, cooling and power required. In plant operation, an energy balance (energy audit) on the plant will show the pattern of energy usage, and suggest areas for conservation and savings.

4.2.2 Conservation of energy

As for material, a general equation can be written for the conservation of energy:

Accumulation = netheat transferredinto the system - net heat transferred out the system-heat consumed +heat generated

This is a statement of the first law of thermodynamics.
An energy balance can be written for any process step.
Energy can exist in many forms and this, to some extent, makes an energy balance more complex than a material balance.
The six important of energy are:

- Work (W) forms of energy that
- Heat (Q) can be transferred
- Kinetic Energy (K)
- Potential Energy (P) forms of energy that
- Internal Energy (U) can be possessed
- Enthalpy (H) by a system

Total energy possessed (E) = K+P+U

Enthalpy H is related to flow systems
Assume that the kinetic and potential energy of the process streams will be small and can be neglected.

Then:
For steady-state processes the accumulation of both mass and energy will be zero**.**
Assume that the kinetic and potential energy of the process streams will be small and can be neglected.
Then:
For steady-state processes the accumulation of both mass and energy will be zero

.

4.2.3 Units operation energy analysis

Rate energy out = Rate energy in

$Qin = Qout$

$Q = mCp\Delta T$ --Eq(4.2)

$Q = mH_l$ -- Eq(4.3)

Where

Q = heat quantity or duty in kj/hr.

m=mass flow rate in kg/hr

cp=specific heat capacity in $kj/kg.c$

Temperature change in C^o.

4.2.3.1Preheater:

285000 kg/hr of nile blent to be heated from $31C^o$ to $198.3C^o$ by exchanging with hot products streams

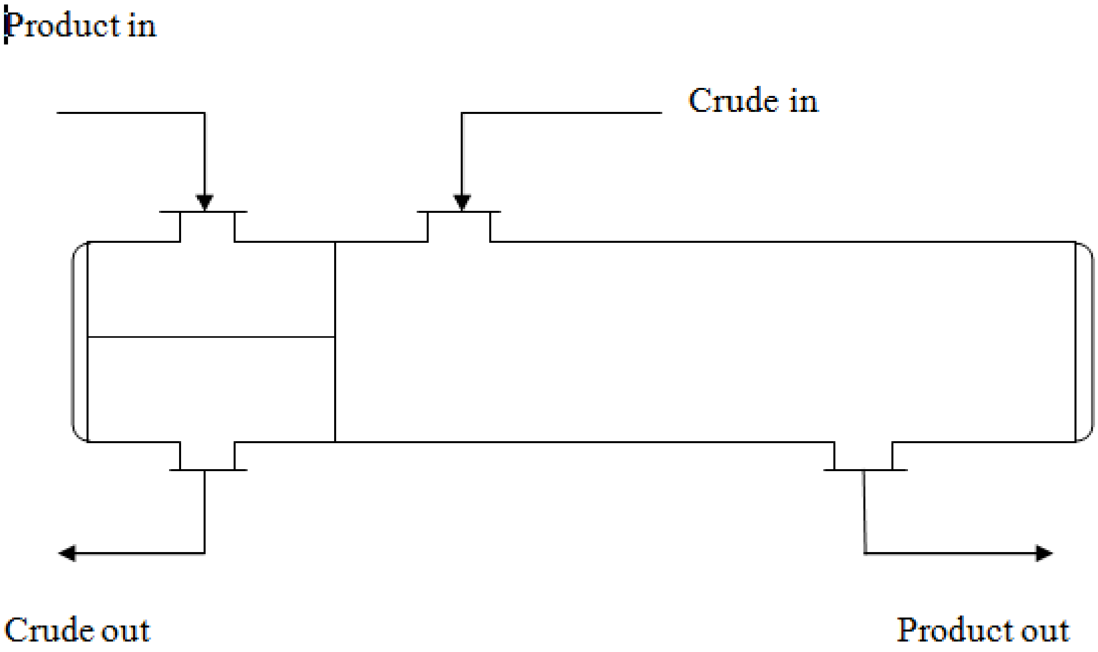


Figure 4.3 heat exchanger input and output streams

First heat exchanger:

Nile Blent with kerosene

Assume steady state operation

$Qgain = Qloss$

Q for kerosene

$m = 2.322 * 10^4\ Kg/hr$

$Cp_{in} = 2.5454\ Kj/Kg.C^o$

$Cp_{in} = 1.985\ Kj/Kg.C^o$

$T_{in} = 160.4C^o \qquad T_{out} = 40.3C^o$

$Cp_{ave} = (Cp_{in} + Cp_{out})/2$

$Cp_{ave} = 2.2652\ Kj/Kg.C^o$

$Q_{loss} = 2.322 * 10^4 * 2.2652 * (160.4 - 40.3) = 6.31 * 10^6 Kj/hr$

Q for Nile Blent

$m = 2.58 * 10^5\ Kg/hr$

$Cp_{in} = 1.87\ Kj/Kg.C^o$

$Cp_{out} = 1.926\ Kj/Kg.C^o$

$T_{in} = 31C^o T_{out} = 43.9C^o$

$Cp_{ave} = 1.898\ Kj/Kg.C^o$

$Q_{gain} = 2.58 * 10^5 * 1.898 * (43.9 - 31) = 6.31 * 10^6 Kj/hr$

For second heat exchanger:

Q for light diesel

$m = 3.406 * 10^4\ Kg/hr$

$Cp_{in} = 2.6087\ Kj/Kg.C^o$

$Cp_{in} = 2.1756\ Kj/Kg.C^o$

$T_{in} = 185.6C^o \qquad T_{out} = 85.3C^o$

$Cp_{ave} = 2.39215\ Kj/Kg.C^o$

$Q_{loss} = 3.406 * 10^4 * 2.39215 * (185.6 - 85.3) = 8.2 * 10^6 Kj/hr$

Q for Nile Blent

$m = 2.58 * 10^5\ Kg/hr$

$Cp_{in} = 1.926\ Kj/Kg.C^o$

$Cp_{out} = 1.9961\ Kj/Kg.C^o$

$T_{in}\ = 43.9C^{o}T_{out}\ = 60.1C^{o}$

$Cp_{ave}\ = 1.96105\ Kj/Kg.C^{o}$

$Q_{gain}\ = 2.58 * 10^{5} * 1.96105 * (60.1 - 43.9) =\ 8.2 * 10^{6}Kj/hr$

For third heat exchanger:

Q for heavy diesel

$m = 1.316 * 10^{4}\ Kg/hr$

$Cp_{in}\ = 2.675Kj/Kg.C^{o}$

$Cp_{in}\ = 2.163\ Kj/Kg.C^{o}$

$T_{in}\ = 212.6C^{o}$ $T_{out}\ = 89.9C^{o}$

$Cp_{ave}\ = 2.419\ Kj/Kg.C^{o}$

$Q_{loss}\ = 1.316 * 10^{4} * 2.419 * (212.6 - 89.9) =\ 3.9 * 10^{6}Kj/hr$

Q for Nile Blent

$m = 2.58 * 10^{5}\ Kg/hr$

$Cp_{in}\ = 1.996\ Kj/Kg.C^{o}$

$Cp_{out}\ = 2.029\ Kj/Kg.C^{o}$

$T_{in}\ = 60.1C^{o}T_{out}\ = 67.3C^{o}$

$Cp_{ave}\ = 2.0125\ Kj/Kg.C^{o}$

$Q_{gain}\ = 2.58 * 10^{5} * 2.0125 * (67.3 - 60.1) =\ 3.9 * 10^{6}Kj/hr$

For forth heat exchanger:

Q for residue

$m = 1.634 * 10^{5}Kg/hr$

$Cp_{in}\ = 2.834Kj/Kg.C^{o}$

$Cp_{in}\ = 2.104\ Kj/Kg.C^{o}$

$T_{in}\ = 295.3C^{o}$ $T_{out}\ = 101C^{o}$

$Cp_{ave}\ = 2.469\ Kj/Kg.C^{o}$

$Q_{loss} = 1.634 * 10^5 * 2.469 * (295.3 - 101) = 7.8 * 10^7 Kj/hr$

Q for Nile Blent

$m = 2.58 * 10^5 \ Kg/hr$

$Cp_{in} = 2.029 \ Kj/Kg.C^o$

$Cp_{out} = 2.563 \ Kj/Kg.C^o$

$T_{in} = 67.3C^o T_{out} = 198.3C^o$

$Cp_{ave} = 2.296 \ Kj/Kg.C^o$

$Q_{gain} = 2.58 * 10^5 * 2.296 * (198.3 - 67.3) = 7.8 * 10^7 Kj/hr$

4.2.3.2 Furnace duty

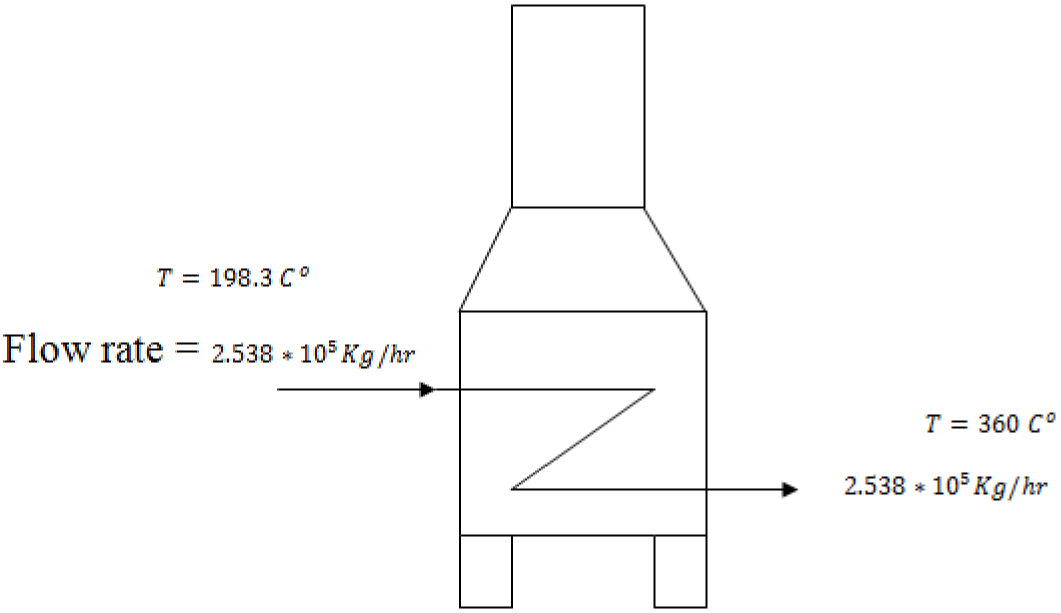


Figure 4.4 Furnace input and output streams

$cp = 3.2 Kj/Kg.C$

$$Q = 2.538 * 10^5 * 3.2 * (360 - 198.3) = 1.31 * 10^8 Kj/hr$$

4.2.3.3 Pumps around

Table 4.3 pump around properties

unit operation	Stream	Temperature (C^o)	Mass flow rate (Kg/hr)	Mass enthalpy (Kj/Kg)
Pump around 1	Draw	117.6	20000	1945
	Return	45.32	20000	2105
Pump around 2	Draw	174.3	75000	1802
	Return	122.8	75000	1930
Pump around 3	Draw	246	40000	1592
	Return	98.01	40000	1966

- Assume all pump around at steady state

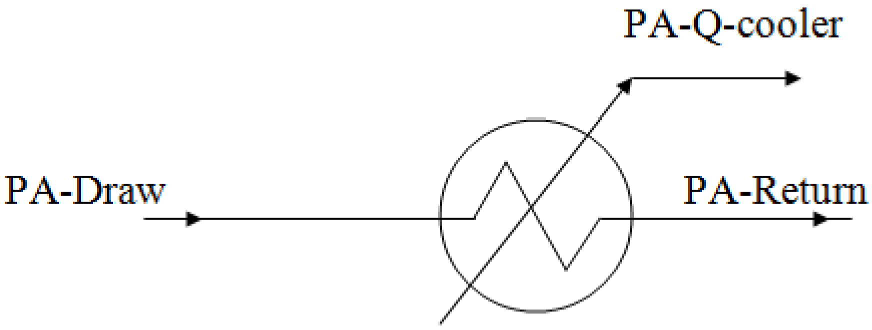


Figure 4.5 Pump around input and output streams

- Pump around 1

Heat of draw steam

$$Q = mH_l$$

$$Q = 2 * 10^4 * 19$$

Heat of return steam

$$Q = mH_l$$

$$Q = 2 * 10^4 * 21$$

Heat removed by pump around 1

$$Q_{in} = Q_{out}$$

$Qdraw = Qremoved + Qreturn$

Q removed = 3.2

- Pump around 2

Heat of draw steam

$Q = 7.5 * 10^4 * 1802 = 1.352 * 10^8 Kj/h$r

Heat of return steam

$Q = 7.5 * 10^4 * 1930 = 1.447 * 10^8 Kj/h$r

Heat removed by pump around 2

$Q = 9.574 * 10^6 Kj/h$r

- Pump around 3

Heat of draw steam

$Q = 4 * 10^4 * 1592 = 6.369 * 10^7 Kj/hr$

Heat of return steam

$Q = 4 * 10^4 * 1966 = 7.866 * 10^7 Kj/hr$

Heat removed by pump around 3

$Q = 1.496 * 10^7 Kj/hr$

4.2.3.4 Condenser duty

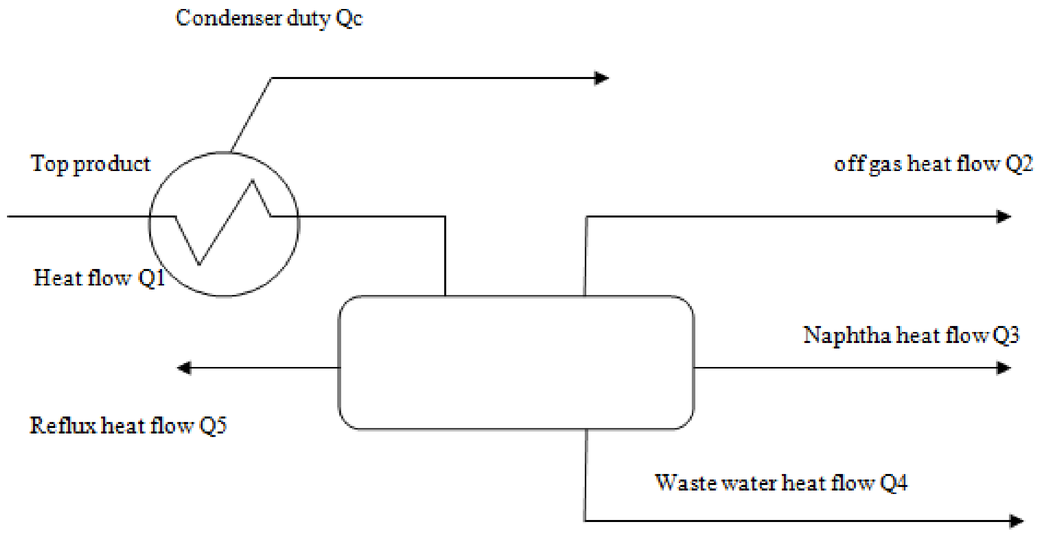


Figure 4.6 Condenser input and output streams

Assume steady state operation

$Qin = Qout$

$Q1 + Qc = Q2 + Q3 + Q4 + Q5$

Table 4.4 heat flow of streams

Stream	Heat flow Kj/hr
Top product	$4.131 * 10^8$
off gas	$1.75 * 10^6$
Naphtha	$5.275 * 10^7$
Waste water	$7.188 * 10^7$
Reflux	$3.809 * 10^8$

$Qc = 9.418 * 10^7 Kj/hr$

4.2.3.5 Reboiler

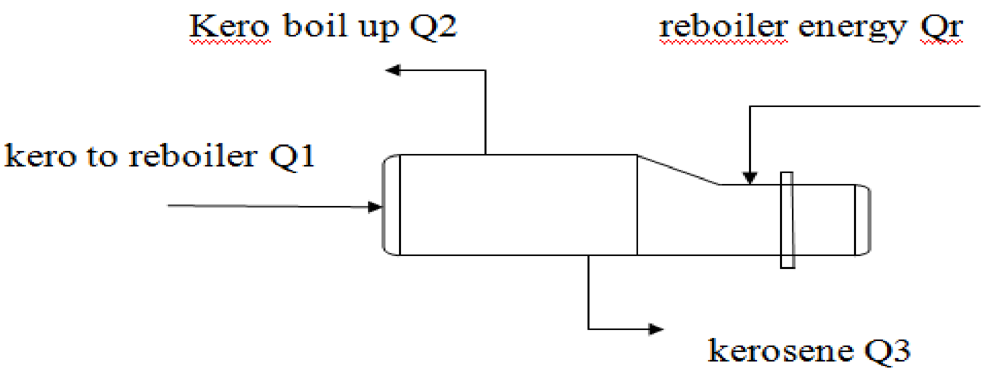


Figure 4.7 Reboiler input and output streams

$Q1 + Qreb = Q2 + Q3$

$Q = m * H_l$

Table 4.5 properties of reboiler streams

Stream	**Mass flow rate** $(\boldsymbol{Kg/hr})$	**Mass enthalpy(** $(\boldsymbol{Kj/Kg})$	**Heat flow** $(\boldsymbol{Kj/hr})$
Kerosene to reboiler	$3.986 * 10^4$	1848.8	$7.369 * 10^7$
Kerosene	$2.322 * 10^4$	1837.3	$4.266 * 10^7$
Kerosene boil up	$1.664 * 10^4$	1577.3	$2.624 * 10^7$

$reb = 4.783 * 10^6 Kj/h$

Table 4.6 duty of units

Unit operation	**Duty** $(\boldsymbol{Kj/hr})$
First heat exchanger	$6.31 * 10^6$
second heat exchanger	$8.2 * 10^6$
third heat exchanger	$3.9 * 10^6$
forth heat exchanger	$7.8 * 10^7$
Furnace	$1.31 * 10^8$
Pump around 1	$3.201 * 10^6$
Pump around 2	$9.574 * 10^6$
Pump around 3	$1.496 * 10^7$
Condenser	$9.418 * 10^7$
Reboiler	$4.783 * 10^6$

4.3 Calculation of Multi-component Distillation Column Design:

The design of a distillation can be divided in the following steps:

1- Select the operating conditions, Operating pressure.

2- Determine the stage and reflux requirement:

- Calculation of Minimum number of stages N_{min}
- Calculation of Minimum Reflux Ratio R_m.
- Calculation of Actual Reflux Ratio.
- Calculation of theoretical number of stages.
- Calculation of actual number of stages

3-Select type of contacting device (Plates or packing).

4- Size the column (Diameter, number of real stages).

5-Design the column internals Plate

6-Total pressure drop over the column
7-Effect of vapor flow conditions on tray design

4.3.1 Input required

- Crude TBP (essential).
- Density/API gravity (essential).
- Molecular Weight(optional).
- Viscosity (optional).

4.3.2 Specification required

- Column Pressure.
- Product specification can be given in terms of fix draw or distillation point.
- Pump around duties need to be specified.
- Column top temperature can be specified

4.3.3 Determine the stage and reflux requirement

4.3.3.1 Selecting the Key components

Table 4.7 heavy and light key

		D	**B**	**K value**
Lk	NBP(O)200*	0.20337	0.000320	2.922
Hk	NBP(O)215*	0.13905	0.000064	2.422

4.3.3.2 Minimum Number of theoretical trays

Using Eq(3.1)

Fenske Equation:

$$N_{min} = \frac{log\ (\frac{xD\ ,l}{xD\ ,H} \times \frac{xB\ ,H}{xB\ ,l})}{\log\ (\alpha l\ ,H)}$$

Where i =LK and j=HK, the minimum number of equilibrium plots is influenced by the components only by their effect on the value of the relative volatility between the key components.
Thus, the minimum number of equilibrium stages depends on the degree of separation of

the two key components and their relative volatility, but is independent of feed-condition

$$(\quad \alpha l, H = \frac{\alpha\ ,lk}{\alpha\ ,Hk} \quad)$$

$$N_m = \frac{log\ (\frac{0.20337}{0.013905} \times \frac{0.000032}{0.000064})}{\log\ (2.922\ /2.422)}$$

$N_m =$ 17.988

4.3.3.3 Minimum reflux from Underwood equation

Minimum reflux from Underwood equation: Minimum reflux is based on the specifications for the degree of separation between two key components. The minimum reflux is finite and feed, product with drawls are permitted.
However, a column cannot operate under this condition because of the requirement of infinite stages. But it is useful limiting condition.
By [using Eq(3.2)]

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q$$

q = condition of feed (heat to vaporized one mole of the feed /molar latent heat of feed).

q = 0.5572

After calculation (using excel)

θ = 1.613676

Underwood equation[from Eq(3.3)]

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_m + 1$$

α_i=the relative volatility of component i with respect some ref. component, usually the
heavy component.
R_m= the minimum reflux ratio.
$X_{i,d}$ = concentration of component i in the tops at minimum reflux

θ=is the root of the equation

$X_{i,f}$= the concentration of component i in the feed.

R_m +1=1.814

R_m = 0.814

R = (1.1 – 1.5) R_m[fromEq(3.4)]

R = 1.2*0.814 = 0.977

Gilliland correlation for actual reflux ratio:

Gilliland (1940) developed an empirical correlation to relate the number of stages N at a finite reflux ratio L/D to the minimum number of stages and to the minimum reflux ratio .

X = $\frac{R-Rm}{R+1}$[from Eq(3.5)]

X= 0.8198

Y = $(1 - X^{1/3})$[from Eq(3.6)]

Y=

$$1-\exp\left[\left(\frac{1+54.4(R-R_{\min})/(R+1)}{11+117.2(R-R_{\min})/(R+1)}\right)\left(\frac{(R-R_{\min})/(R+1)-1}{[(R-R_{\min})/(R+1)]^{0.5}}\right)\right]$$ [from Eq(3.7)]

Y=0.565

$N = \frac{N_{min} + Y}{1-Y}$[from Eq(3.8)]

$= 42.6\ stages$

From figure 3.1

R/(R+1) =0.494

R_m/ (R_m +1) =0.448

N/N_m=0.425

N=42.6

4.3.3.4 Calculation of actual number of stages

4.3.3.4.1 Estimating Efficiencies – The O'Connell Method

There are many methods that have been developed to estimate distillation efficiencies. Here we consider just one method; that of H.E. O'Connell (*Trans. AIChE*, **42**, 741, 1946). O'Connell obtained his correlation for the efficiency of distillation processes from an analysis of data on several operating columns. The original correlation was graphical, but equations have been proposed to represent the correlation. One such equation is:

$E_{oc} = 50.3(\alpha\mu)^{-0.226}$[fromEq(3.9)]

Where is α the relative volatility between the key components and μ is the viscosity in cp.

E_{oc} = $50.3(1.2 * 0.1139)^{-0.226}$
E_{oc} = 79.18 %

By using the figure 3.2

From the graph is 79.4%

Actual number of stages = $\frac{\text{N theoritical}}{\text{E}}$ [from Eq(3.10)]

Actual number of stages = $\frac{42.59}{0.7918}$

Actual no of stages =53.6 Stages

$Efficency = 0.17 - 0.616 log_{10}$ ⍰------------------Eq(5.1)-

=75.118%

4.3.4 Feed plate location

The feed location is determined by the Kirkbridge equation

$$\log\left(\frac{N_R}{N_S}\right) = 0.206 \log\left[\left(\frac{D}{W}\right)\left(\frac{x_{f,HK}}{x_{f,LK}}\right)\left(\frac{x_{b,LK}}{x_{d,HK}}\right)^2\right]$$ ----------Eq (3.11)

D=145 Kmole/hr

B=397.5 Kmole/hr

$X_{f,HK}$= 0.028158

$X_{f,LK}$=0.032925

$X_{b,LK}$=0.00007

$X_{d,HK}$=0.013905

$\frac{N_r}{N_s}$=0.134717

N_s=N_T (1+ N_r/N_s)=6.28 stages

4.3.5Column daimeter

The principal factor that determines the column diameter is the vapour flow-rate. Thevapour velocity must be below that which would cause excessive liquid entrainment or ahigh-pressure drop. The equation given below, which is based on the well-known Souders and Brown equation, Lowenstein (1961), can be used to estimate the maximum allowable superficial vapour velocity, and hence the column area and diameter.

$$\hat{u}_v = (-0.171 l_t^2 + 0.27 l_t - 0.047)\left[\frac{(\rho_L - \rho_v)}{\rho_v}\right]^{1/2}$$ [ftomEq (3.12)]

Where:

$\hat{u}_v$maximum allowable vapour velocity, based on the gross (total) column

cross-sectional area, m/s,

ι_t = plate spacing, m, (range 0.5 1.5).

t = tray spacing (m), assume it 0.5 .

Vw = mass vapor flow rate = 72580 kg/h.

Lw = mass liquid flow rate = 185400 kg/h.

ρ_v = Vapour density =15.19 kg/m^3.

ρ_L = Liquid density =617.1 kg/m^3.

.$\hat{u}_v = 0.2846225\, m/s$

The column diameter D_c, can then be calculated:

$$D_c = \sqrt{\frac{4V_W}{\pi\rho_v \hat{u}_v}}$$ [fromEq (3.13)]

whereVw is the maximum vapour rate, kg/s.

D_c= 2.4 m.

4.3.6 Height of Distillation Column

Height of column

$H_c = (N_{act} - 1)\ H_s + \Delta H +$ plates thickness[from Eq (4.3)]

No. of plates = 53.6 trays

Tray Spacing: H_s = 0.46 to 0.61m.

H_s = 0.5 m.

ΔH= 0.5 meter each for liquid hold up and vapor disengagement

ΔH=1 m.

Total thickness of trays = 0.005*53.6= 0.268 m.

Height of column = (53.6 - 1)*0.50+ 1+0.259= 27.6m.

4.3.7 Plate layout Design

Platetype : Sieve .

Column diameter = 2.4m

A_t = tower cross sectional area m^2.

=4.66m^2

Flow Parameter4.3.7.1

$$F_{LV} = \left(\frac{L_n}{V_n}\right)\left(\frac{\rho_V}{\rho_L}\right)^{0.5} \text{---} \text{Eq (4.4)}$$

$$= 0.40077$$

$$At = \frac{An}{1-rd} \qquad \text{---Eq (4.5)}$$

$$r_d = A_d/A_t \qquad \text{---Eq (4.6)}$$

Table 4.8 relation between F_{LV} and r_d

≤0.1	**0.1**
1-0.1	**0**
≥1	**0.2**

$r_d = 0.1+\frac{0.40077-0.1}{9}$

$= 0.1334m.$

$$4.66 = \frac{An}{1-0.1334}$$

A_n= net area of the tray m^2

$= 4.038m^2.$

A_d=down comer area m^2

$A_d = A_t * 0.12$ ------------------------------------ Eq(4.7)

$= 0.12*4.66=0.5592m^2$

A_a = active aream^2

$A_a = A_t - 2*A_d$ -------------------------------------Eq (4.8)

= 4.66-2*.5592

$= 3.5416\ m^2$

Hole area m^2

$A_h = 0.10*A_a$ ---------------------------------------Eq (4.9)

= 0.10*3.5416

$=0.35416m^2$

Hole diameter should be in the range of (3.2-12.7)mm

Assume hole diameter D_o =8mm

Tray thichness= (0.65-0.15)D_o=0.63*8=5.04mm

Area of single hole =((3.14)*(8*10^-3)^2)/4 =0.00005m^2

$No = \frac{4*Ah}{pi*do^2}$--Eq (4.10)

= 7050 holes

Hole pitch L_P =2.5*8=20mm

5.7.2 Weir length

$L_w = \frac{Ad}{At}$--Eq (4.11)

= 0.5592/4.66=0.12000

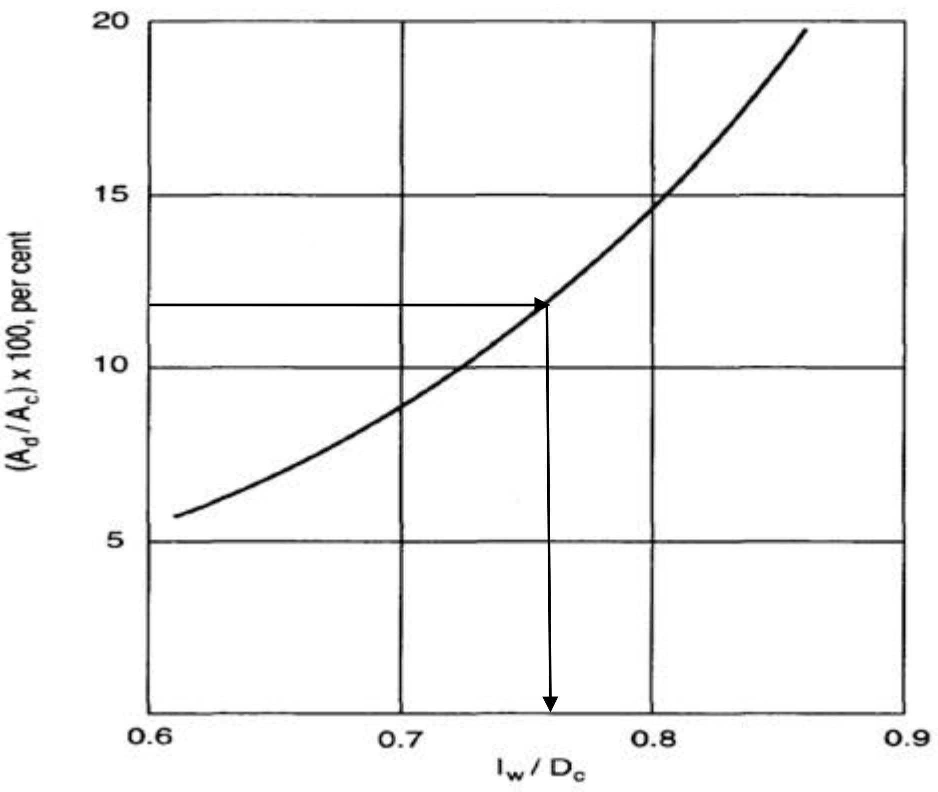


Figure 4.8Relation between downcomer area and weir length

L_w/D_t=0.76

L_w =2.4*0.76=1.824m

Weir length should be 60 to 85% of column diameter

4.3.8 Pressure drop through the tray

4.3.8.1 Dry plate drop hd

Dry plate pressure drop occurs due to friction within dry short holes.hdcan be calculated using following expression derived for flow through orifices

$h_d = 51[\frac{u_h}{C_o}]^2 * \frac{\rho_v}{\rho_L}$ [fromEq(3.14)]

u_h= vapor velocity through holes m/s

$u_h = \frac{qv}{Ah}$ [from Eq (3.15)]

$=\frac{1.32901194}{0.221844}$ =5.99074 m/s

C_o=orifice coefficient (prince correlation)

C_o=$(1.625 - 0.625 * (A_h/A_a))^{-.05}$ ---------------------------Eq (4.12)

A_h/A_a = $\frac{0.221844}{2.21844}$ =0.1

Co=0.8

H_d=0.0703 $m.$

- H_w = weir height

Assumed (50mm=0.05 m)

- H_{ow} =liquid crestover weir

$h_{ow} = 750[\frac{L_w}{\rho_L \iota_w}]^{2/3}$--------------------------Eq(4.13)

ρ_L (max liquid flow rate)

= 0.08345 $m^3/s.$

Min ρ_L (at 70 percent turn-down)=0.7*.08345=0.058 $m^3/s.$

H_{ow} =74.41$mm.$

(Foss and crester , 1956)

H_L=Fa*(H_w+H_{ow}) ----------------------------------Eq (4.14)

Fa=aeration fraction (0.6-1)(assumed 0.20)

H_L= 0.03266 m

- h_r= residual pressure drop

The residual pressure drop results mainly from the surface tension as the gas releases from a perforation. The following simple equation can be used to estimate $hr$with reasonable accuracy

$h_r = \frac{12.5*10^3}{\rho L}$ --------------------------------Eq (4.15)

=0.02025m

Total pressure drop per plate is then cumulatively found from:

$h_t(m) = h_d + (h_w + h_{ow}) + h_r$[fromEq(3.17)]

Total pressure drop over entire column may be then calculated from the formula :

$\Delta P_{total} = (9.81 * 10^{-3}) * (H_t) * (\rho_{L)}$ ------------------------------------Eq(4.16)

$$Ht = 0.07039 + 0.2(0.05 + 0.0744) + 0.02025 = 0.11252m$$

Pressure drop=0.6993 kpa =0.1014 psi

4.3.9 Effect of vapor flow conditions on tray design

4.3.9.1 Flooding check

Volumetric flow rate of vapor = Q_v

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

mass vapor flow rate = 72580 kg/h

Q_v=1.32901194 m^3/s

Now, net area

$V_n = \frac{\text{Qv}}{\text{An}} == \frac{1.32901194}{4.038}$ = 0.329 m/s

The flooding velocity can be estimated from the correlation given by Fair (1961):

$$u_f = K1\sqrt{\frac{\rho_L - \rho_V}{\rho_L}}$$ [fromEq (3.18)]

u_f flooding vapour velocity, m/s, based on the net column cross-sectional areaA_n.

K1 a constant obtained from Figure 3.3

K1=0.059

u_f = 0.059*6.29=0.37111m/s

u_f >U_n(u_n = 0.88*u_f)

No flooding occur

4.3.9.2 Check of weeping

H_w+H_{ow} =124.4mm

From figure 3.4

K2=31.2

U_{min} at weep point = 2.141716m/s

Actual operating minimum vapor velocity:

$$U_{\min,\text{op}} = \frac{minimum\ \ vapor\ \ flow\ rate}{hole\ area}$$ [fromEq (3.19)]

= 1.329/0.35416 = 3.7526 m/s

3.7526 m/s>2.41716 m/s

To avoid weeping$U_{min\ .op} > U_{min}$.

Therefore Weeping will not take place.

4.3.9.3 Check of fractional entrainment

Entrainment is the phenomena in which liquid droplets are carried by vapor/gas to the tray above Entrainment can be estimated from the correlation given by Fair (1961), Figure 11.29, which gives the fractional entrainment ψ (kg/kg gross liquid flow) as function of the liquid-vapour factor FLV, with the percentage approach to flooding as a parameter. The percentage flooding is given by:

Percent flooding = U_n/ U_f

%flooding (Fp)= 0.5253/0.6168=0.85=85%

F_{LV} = 0.40077

From the figure 3.5 we observed that

Ψ = 0.09

ifψ ≤0.2 the design is ok

4.3.9.4 Check of residence time in the down comer

L_d = H_w+C

L_d=0.05+0.5=0.55m

= 0.03328/0.3503=0.095m/s

Ţ=L_d/U_d = 5.789s

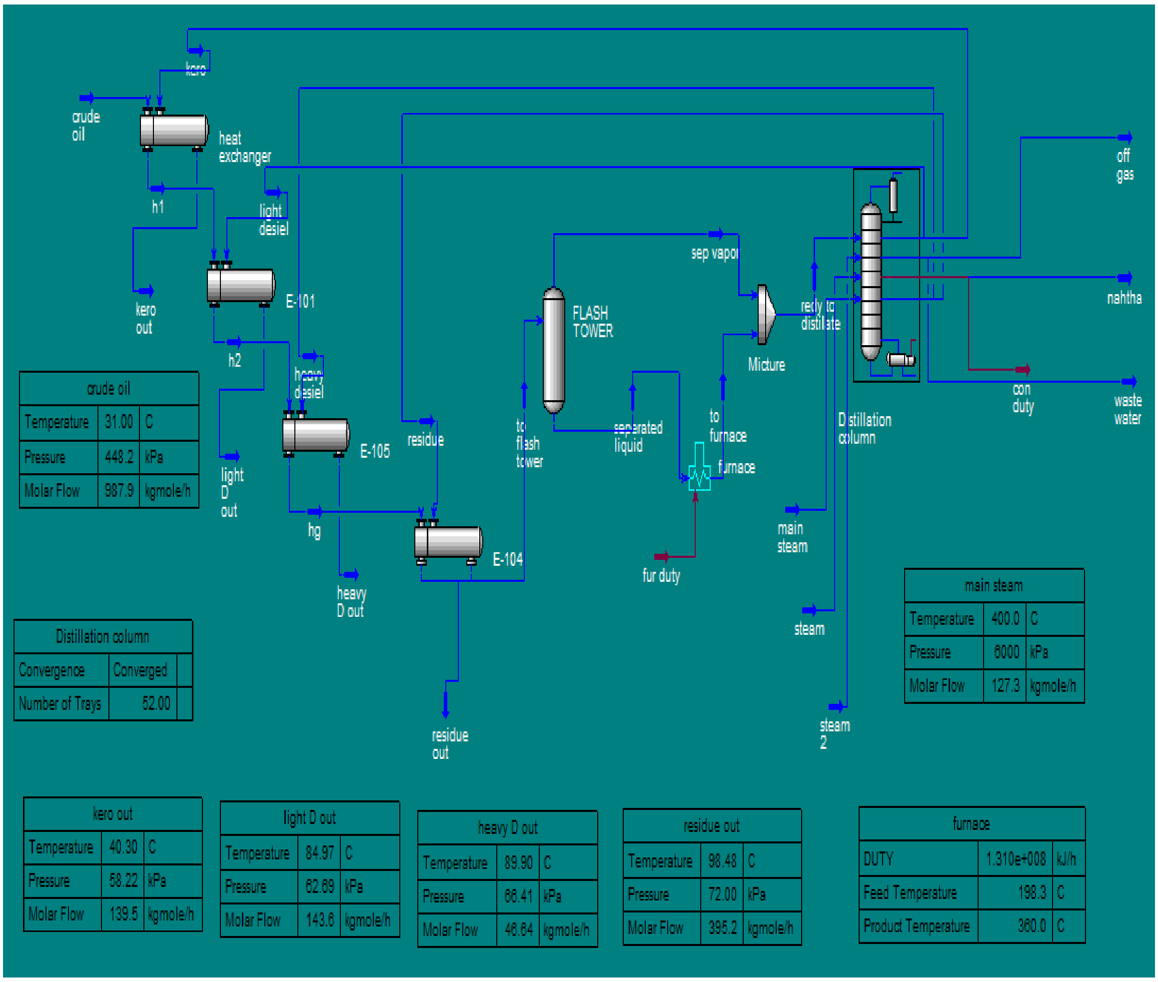
If Ţ>3s , ok

4.4 summary of results:

4.4.1 Design outcomes

Table 4.9 Design specification of atmospheric crude distillation column based on the Nile blend assay 2004

No	**Items**	**Specification**	**Unit**
Column diameter	D_c	**2.4**	m
Column height	H_c	**27.6**	m
Min reflux ratio	R_{min}	**0.814**	**-**
Actual reflux ratio	R	**0.977**	**-**
Feed condition	q	**0.5572**	**-**
Efficiency	E	**79.18**	**%**
Flow parameter	F_{LV}	**0.40077**	**-**
Total area	A_t	**4.66**	m^2
Net area	A_n	**4.038**	m^2
Dawncomer area	A_d	**0.5592**	m^2
Holes area	A_h	**0.35416**	m^2
No of holes	N_o	**7050**	**Hole**
Hole Pitch	L_p	**20**	mm
Tray space	C	**0.5**	m
Tray thickness	Y	**5.04**	m
Weir length	L_w	**1.8**	m
Feed location	N_s	**6**	**-**
Weir height	H_w	**0.5**	m
Pressure drop	$P1 - P2$	**0.1014**	psi
Fractional entrainment	Ψ	**0.09**	**-**
Residence time	**T**	**5.8**	s
Number of trays	**N**	**53.6**	**-**



crude oil		
Temperature	31.00	C
Pressure	448.2	kPa
Molar Flow	987.9	kgmole/h

Distillation column		
Convergence	Converged	
Number of Trays	52.00	

main steam		
Temperature	400.0	C
Pressure	6000	kPa
Molar Flow	127.3	kgmole/h

kero out		
Temperature	40.30	C
Pressure	58.22	kPa
Molar Flow	139.5	kgmole/h

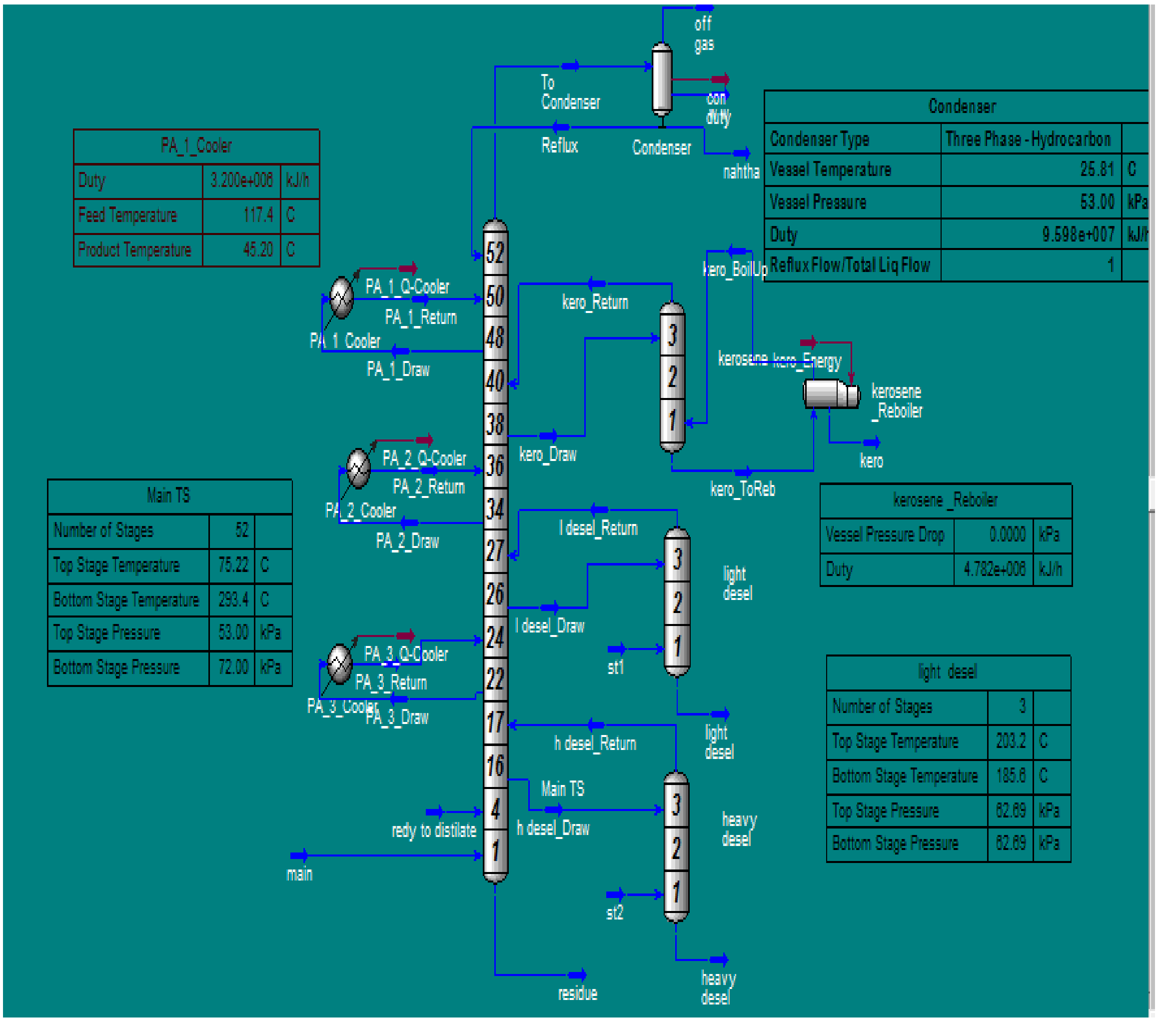
light D out		
Temperature	84.97	C
Pressure	62.69	kPa
Molar Flow	143.6	kgmole/h

heavy D out		
Temperature	89.90	C
Pressure	66.41	kPa
Molar Flow	46.64	kgmole/h

residue out		
Temperature	98.48	C
Pressure	72.00	kPa
Molar Flow	395.2	kgmole/h

furnace		
DUTY	1.310e+008	kJ/h
Feed Temperature	198.3	C
Product Temperature	360.0	C

Figure 4.9 Simulation process flow sheet result.



PA_1_Cooler		
Duty	3.200e+006	kJ/h
Feed Temperature	117.4	C
Product Temperature	45.20	C

Condenser		
Condenser Type	Three Phase - Hydrocarbon	
Vessel Temperature	25.81	C
Vessel Pressure	53.00	kPa
Duty	9.598e+007	kJ/
Reflux Flow/Total Liq Flow	1	

Main TS		
Number of Stages	52	
Top Stage Temperature	75.22	C
Bottom Stage Temperature	293.4	C
Top Stage Pressure	53.00	kPa
Bottom Stage Pressure	72.00	kPa

kerosene _Reboiler		
Vessel Pressure Drop	0.0000	kPa
Duty	4.782e+006	kJ/h

light desel		
Number of Stages	3	
Top Stage Temperature	203.2	C
Bottom Stage Temperature	185.6	C
Top Stage Pressure	62.69	kPa
Bottom Stage Pressure	62.69	kPa

Figure4.10 column environment results.

4.4.2 Column profile

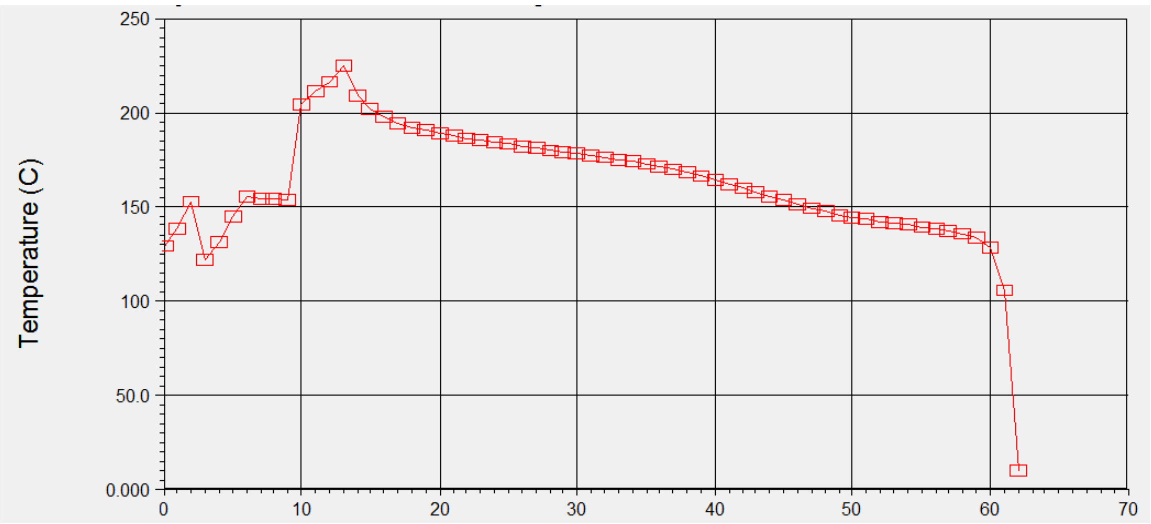


Figure 4.11 Temperature vs Tray position from bottom

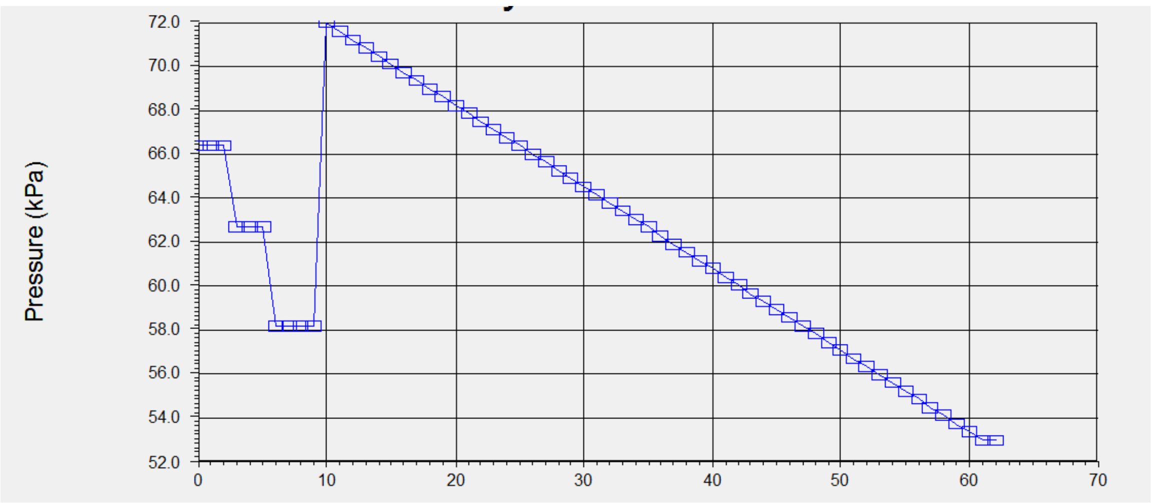


Figure 4.12 pressure vs Tray position from bottom

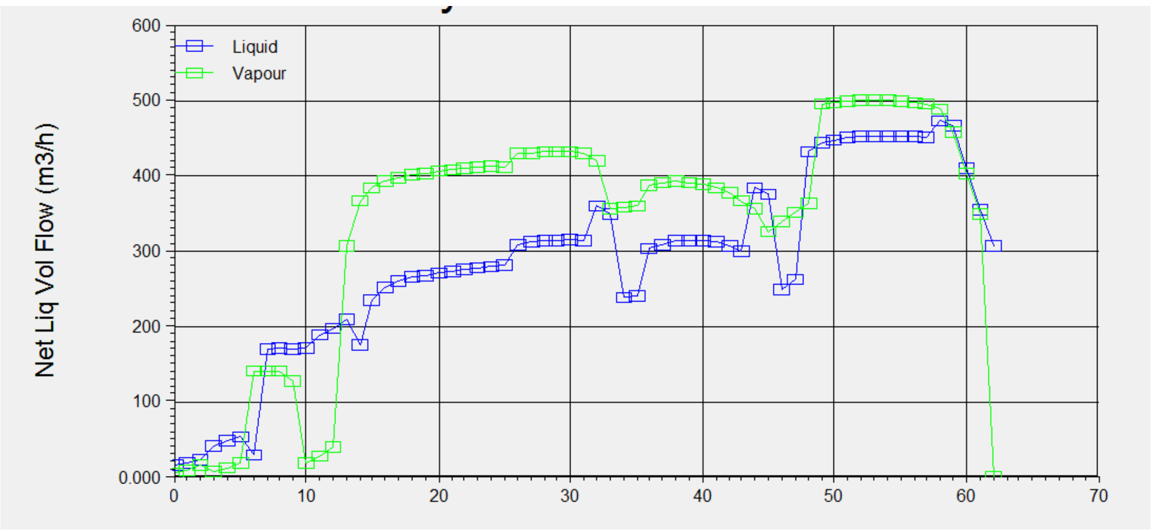


Figure 4.13Net liquid flow vs Tray position from bottom

4.4.3 Properties variation among different crudes (6 samples)

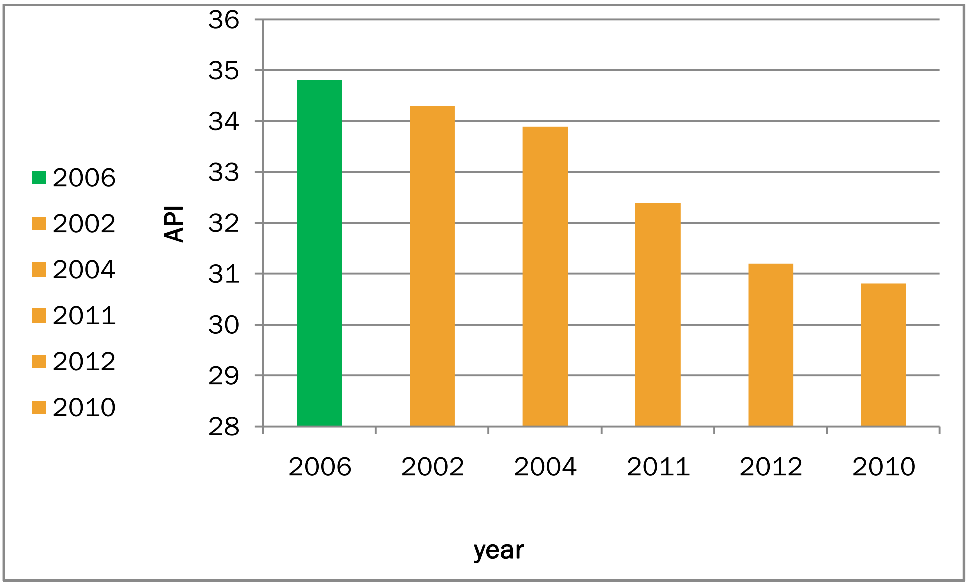


Figure 4.14 API variation of different crudes assay.

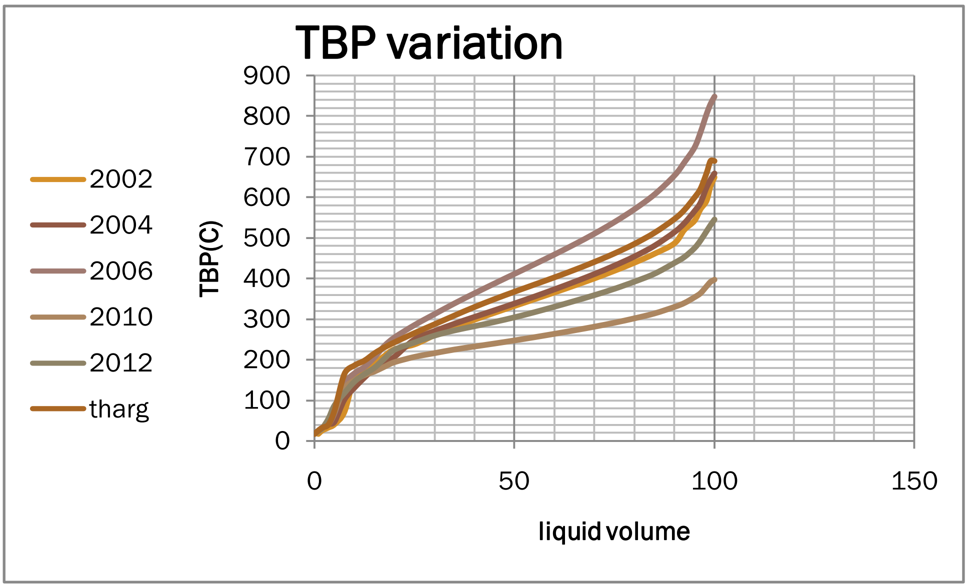


Figure 4.15 Density variation of different crudes assay

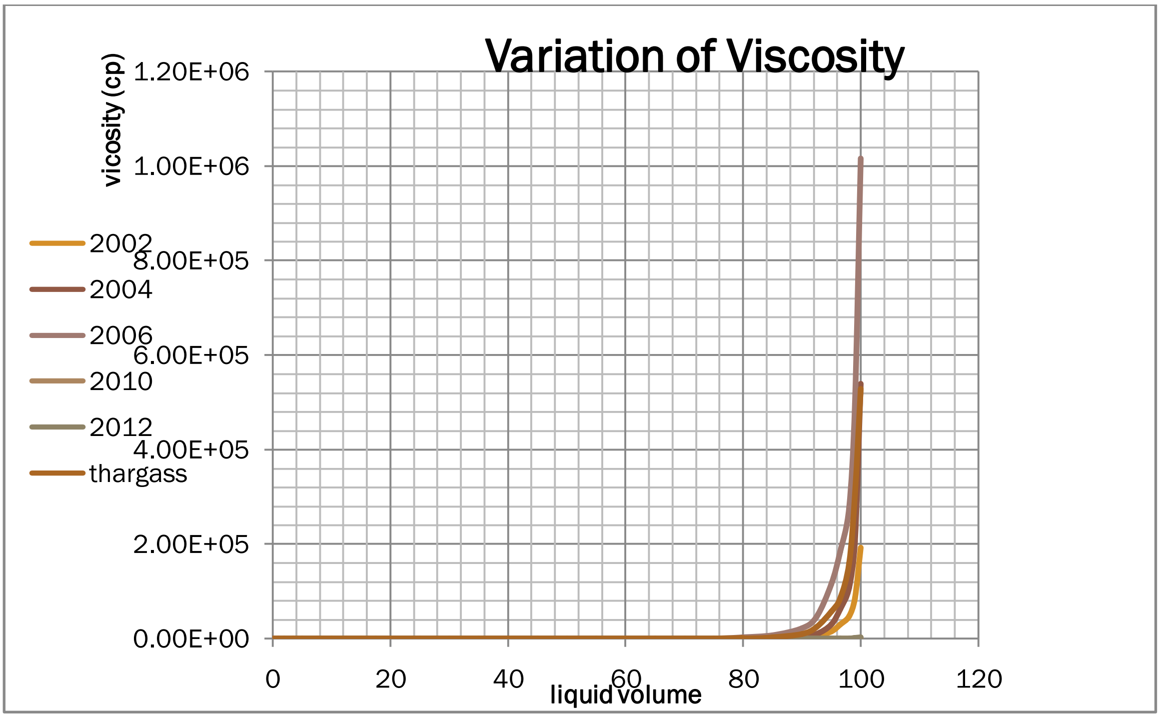


Figure 4.16Viscosity variation of different crudes assay.

4.4.4 Comparison of distillation tower products :

We simulate different crude assay using HYSIS as simulator tool and the variation in the amount of atmospheric distillation product as primary result are shown in the table bellow:-

Table 4.10 products quantities simulation results.

No	**Crude name**	**Date**	**API**	**naph**	**Kero**	**L D**	**H D**	**Res**
1	**Nile blend**	**2006**	**34.81**	**24.51**	**24.51**	**32.25**	**16.4**	**159.98**
2	**Nile blend**	**2002**	**34.3**	**23.736**	**25.025**	**33.798**	**12.9**	**162.28**
3	**Nile blend**	**2004**	**33.9**	**23.478**	**23.22**	**34.056**	**13.158**	**163.31**
4	**Thargas**	**-**	**32.4**	**20.124**	**21.156**	**32.766**	**12.384**	**171.31**
5	**Nile blend +LCO**	**2012**	**31.2**	**13.416**	**16.254**	**26**	**20.182**	**182.15**
6	**Nile blend**	**2010**	**30.81**	**12.332**	**9.5976**	**10.446**	**9.446**	**216**

\

Table 4.11 Atmospheric Residue and naphtha percent in the different runs

API	Res %	Naph%
34.81	62.007	9.5
34.3	**62.9**	**9.2**
33.9	**63.3**	**9.1**
32.4	**66.27**	**7.8**
31.2	**70.5**	**5.2**
30.81	**83.7**	**4.76**

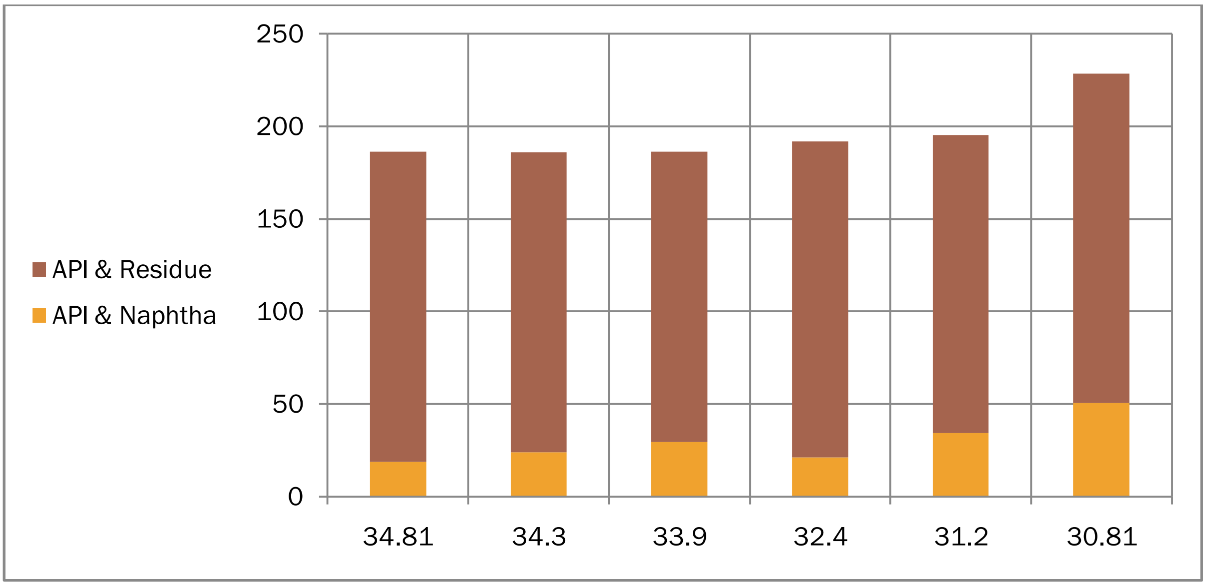


Figure 4.17variation of top and bottom product percent according to API

Table 4.12 the deviation on design variables with the change of assay .

Items	Actual design of KRC	Design based on 2004 assay	Design based on 2006 assay
Number of stages	**52**	**53.6**	**51.8**
Column diameter(m)	**2**	**2.4**	**2**
Column height(m)	**26**	**27.6**	**26.6**

4.4.5The effect of reflux ratio on the bottom and top products:

Table 4.13 Fixing the reflux ratio to be 7.5 for all crudes (6 samples) we observe the following change in the naphtha and residue

API	naphtha	Naphtha@7.5	residue@7.5	Residue
34.81	24.51	13.7	159.98	173
34.3	23.736	17.7	162.28	168
33.9	23.478	22.4	163.31	164
32.4	20.124	15.6	171.31	176
31.2	13.416	25	182.15	171
30.8	**12.332**	36.6	216	192

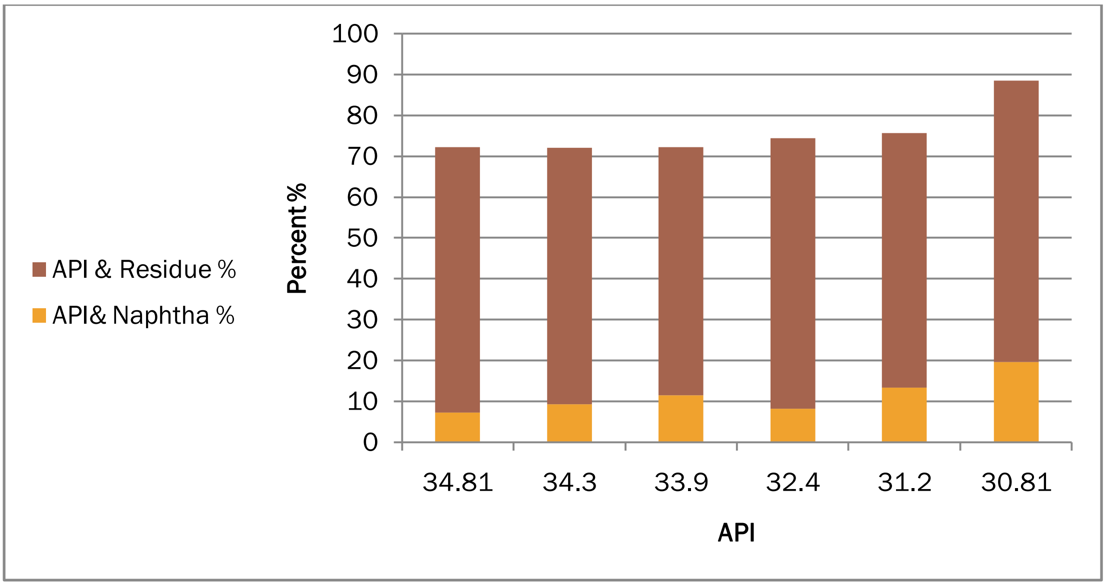


Figure 4.18 productivity change of top and bottom with changing reflux ratio to 7.5

Table 4.14 Fixing the reflux ratio to be 5 for all crudes (6 samples) we observe the following change in the naphtha and residue

API	Naphtha	naphtha@ 5	Residue@5	Resiue
34.81	24.51	18.5	168	173
34.3	23.736	24	162	168
33.9	23.478	29.5	157	164
32.4	20.124	21	171	176
31.2	13.416	34.2	161	171
30.8	**12.332**	50.5	178	192

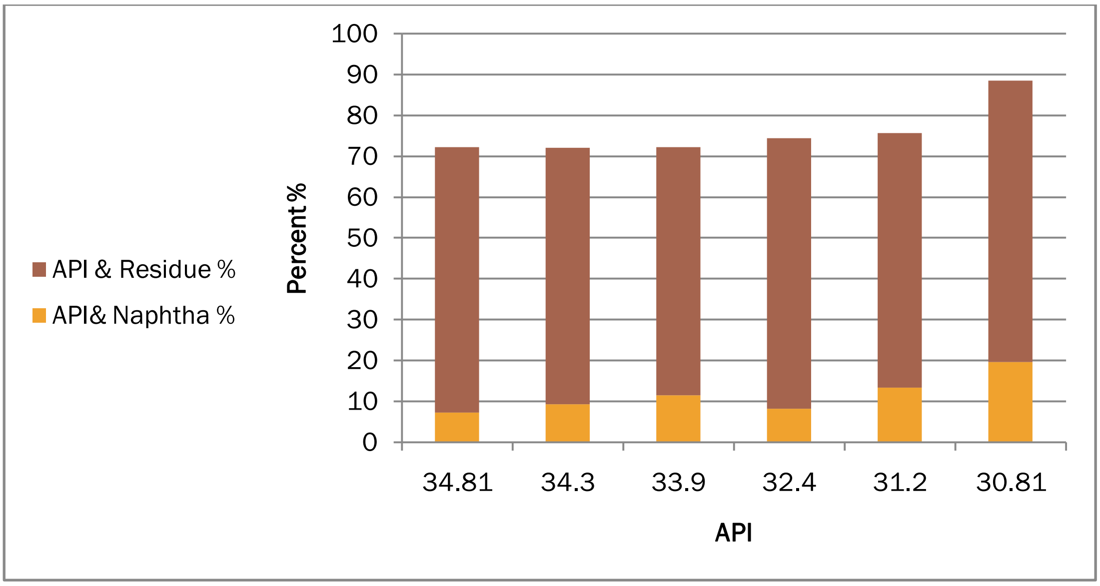


Figure 4.19 productivity change of top and bottom with changing reflux ratio to 5

Table 4.15 the effect of reflux ratio on Nile blend crude assay 2010top and bottom productivity .

Reflux	26.8	20	8.5	6	5
Naphtha %	4.7	11.05	12.59	17.05	19.5
Residue%	83.7	77.52	75.96	71.705	68.98

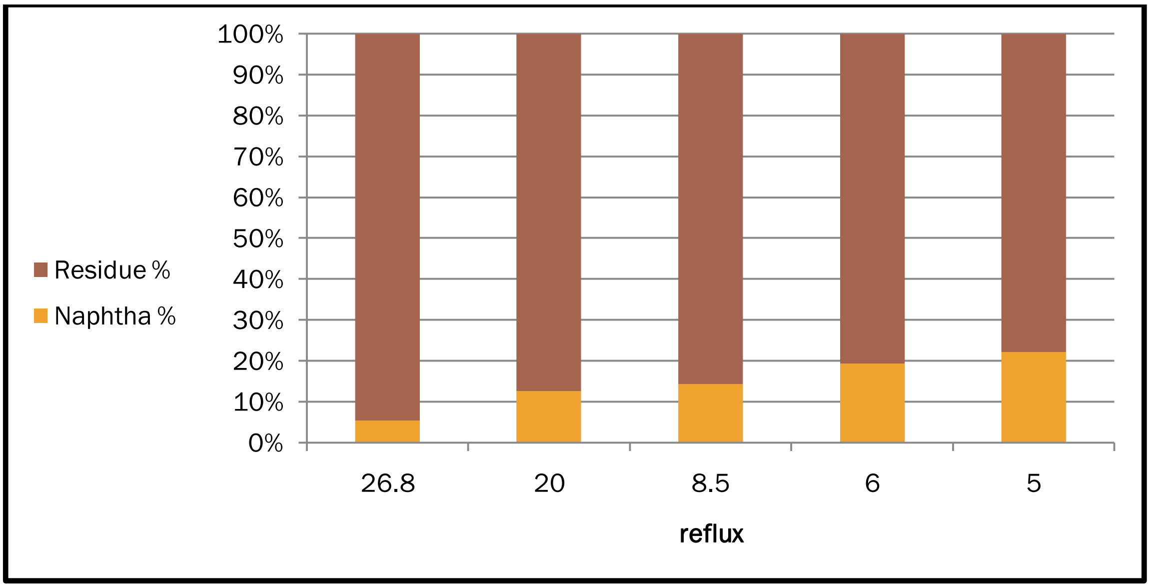


Figure 4.20 the effect of reflux ratio on Nile blend crude assay 2010top and bottom productivity .

4.4.6 Multi feed effect on naphtha productivity

Table 4.16Instead of entering the feed on the third plate (one location) the feed has been divided (30%,30%,40%) and enter to the tower at different location (8,5,3 respectively) fixing the other parameters constant:

Feed location	Naphtha (ton/h)	Residue (ton/h)	Naphtha%	Residue %
One feed location	23.5	164	9.1	63.5
Three feed location	30.2	157	11.78	60.9
		%change	2.68	2.6

Discussion:

From 2002 to 2012 The differentiation of API value during this time from 34.81 to 28.8 and this reflect the change in crude oil properties , and based on that detailed study to identify the effect of this variation , the primary comparison present that the quantity of naphtha from 2002 to 2012 has decreased approximately to half of it is initial value , and increased the amount of residue from 62% to 83% and this point present a capability to increase the profitability as economic factor , using multi feed location as well as reflux ratio has a significant impact in enhancing the productivity of naphtha , but changing the number of trays and pumps around flow rates does not has considerable effect .

The 2012 crude oil has API° is 28.8, Sulfur content is 0.072 m%, salt content is 3.5 mg/l as NaCl and these properties does not match the distillation column specification , there is a necessary to search for a better method to handle with this crude , the practical method which is actually occur and applied in the KRC is blending the crude oil 2012 with light crude oil which is LCO the blending ratio is 70 percent Nile blend crude oil with 30 percent LCO Crude Oil . The LCO crude came from Petro Energy and Nile blend crude oil sampled from the Sudan Wells , or separation the crude oil by reducing the pressure on many stage according to cuts range temperature .

4.5Cost estimation Calculation

4.5.1Equipment cost

4.5.1.1-direct-fierd heater:

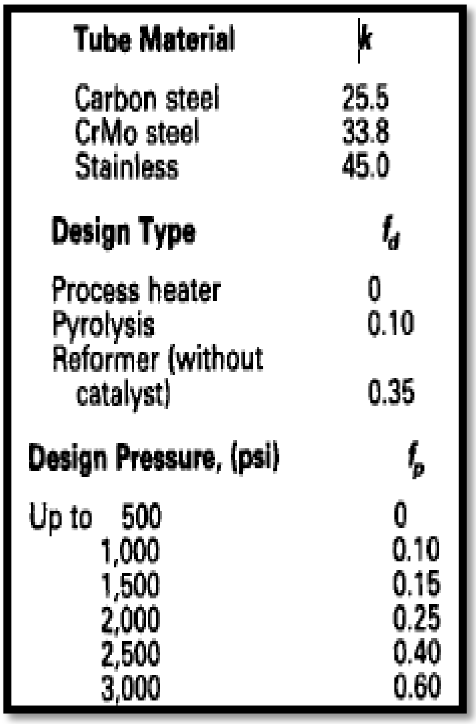
$C_s, \$=k*(1+F_d+F_p)*Q^{0.86}$ ---.Eq(4.16)

Where :

20<Q>200 MBtu/hr.

Cs = cost of direct-fired heater $

Table (4.17) Correction Factor of direct-fired heater



Tube Material	k
Carbon steel	25.5
CrMo steel	33.8
Stainless	45.0
Design Type	f_d
Process heater	0
Pyrolysis	0.10
Reformer (without catalyst)	0.35
Design Pressure, (psi)	f_p
Up to 500	0
1,000	0.10
1,500	0.15
2,000	0.25
2,500	0.40
3,000	0.60

K = tube material

F_d =design type

Design pressure, (psi)

From data

Q=1.24207*10^2 MBtu/hr

K=45(stainless).

F_d =0 (process heater).

F_p =0.6(3000 psi design pressure)

C_s=45*(1+0+0.6)*(1.24207*10^2)^0.86

C_s=4553 $

From (Table C) the installation factor is 2.1 so that the installed
Price is = 0

Installation cost=45553*1.3

=59218.9$

4.5.1.2 Cost of distillation column :

C_t, \$=F_1*C_p+N*F_2*F_3*F_4*C_{tray}+C_{bl}--Eq (4.17)

C_p=exp [7.123+0.1478(ln W)+0.02488*(ln $W)^2$ +0.01580*($\frac{L}{D}$)+ln $\frac{Tb}{Tp}$]------Eq (4.18)

9020<w>2,470,000Ibs

C_{tray} =375.8*exp [0.1739*D]---Eq(4.19)

2<D>16ft

N=number of trays.

C_{bl}=204.9*$D^{0.6332}$ *$L^{0.8018}$ --Eq(4.20)

2<D>20ft , 25<L>170ft

T_b==thickness of shell at the bottom (0.5in).

T_p=thickness required for operating pressure (0.75).

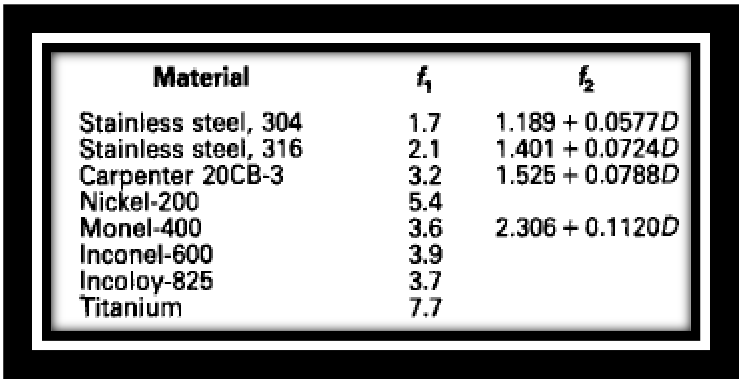
F_4=$\frac{2.25}{1.0414^N}$--.Eq(4.21)

W=$\frac{\pi}{4} * \left(\frac{16 * wallt\ hckness\ * 501}{12}\right) + 2 * 325$--------------------------------------Eq(4.22)

(325) flanged and dished heads weigh 325 Ib each

W=24391.3Ibs

Table 4.18Material factors of column



Material	f_1	f_2
Stainless steel, 304	1.7	$1.189 + 0.0577D$
Stainless steel, 316	2.1	$1.401 + 0.0724D$
Carpenter 20CB-3	3.2	$1.525 + 0.0788D$
Nickel-200	5.4	
Monel-400	3.6	$2.306 + 0.1120D$
Inconel-600	3.9	
Incoloy-825	3.7	
Titanium	7.7	

F_2=1.401+0.0724*7.874=1.97(stainless steel 316)

F_3=1.59(bubble cap tray)

F_4=2.25/(1.0414)^52

C_p=exp [7.123+0.1478*(Ln24391.3)+0.02488(Ln24391.3)^2+0.0158(90.55/7.874)*Ln(0.5/0.75)]

C_p=64952.11$

C_{tray} =375.8*exp[0.1739*7.874)

C_{tray} = 1477.85$

C_{bl}= 204.9*7.874^0.6332*90.55^0.8018

C_{bl} =28057.3$

C_t=2.1*64952.11+52*1.97*1.59*2.25*1477.8+28057.3

C_t=706039.54$

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

Installation cost, (C_{int})=706039.54*2.1

C_{int} =1,482,683.044

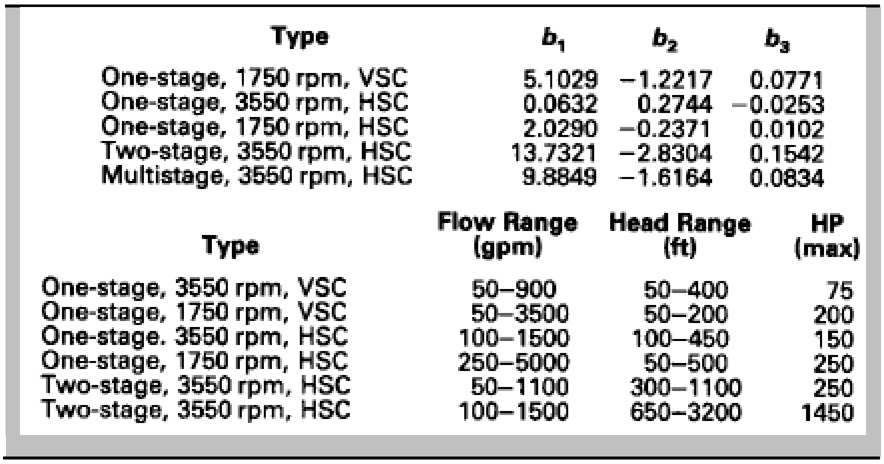
4.5.1.3 Cost of pumps:

C= F_M*F_D*C_p--.Eq(4.23)

C_p=1.55*exp [8.833-0.6019*ln($Q * \sqrt{H}$) +0.0519*(ln($Q * \sqrt{H}$))^2]------Eq(4.24)

Materiel	F_M
Stainless steel	2.00

Table 4.19 Factor of type & Head Range Of Pump



Type	b_1	b_2	b_3
One-stage, 1750 rpm, VSC	5.1029	−1.2217	0.0771
One-stage, 3550 rpm, HSC	0.0632	0.2744	−0.0253
One-stage, 1750 rpm, HSC	2.0290	−0.2371	0.0102
Two-stage, 3550 rpm, HSC	13.7321	−2.8304	0.1542
Multistage, 3550 rpm, HSC	9.8849	−1.6164	0.0834

Type	Flow Range (gpm)	Head Range (ft)	HP (max)
One-stage, 3550 rpm, VSC	50–900	50–400	75
One-stage, 1750 rpm, VSC	50–3500	50–200	200
One-stage. 3550 rpm, HSC	100–1500	100–450	150
One-stage, 1750 rpm, HSC	250–5000	50–500	250
Two-stage, 3550 rpm, HSC	50–1100	300–1100	250
Two-stage, 3550 rpm, HSC	100–1500	650–3200	1450

F_M=exp [b_1+b_2*ln($Q*\sqrt{H}$) +b_3*(ln($Q*\sqrt{H}$))^2]-----------------.Eq (4.25)

Cost of each pump

F_t =exp [9.8849-1.6164*Ln333.33*sq250+0.0834*(Ln333.33*sq250)^2]

F_t=8.65F_m=2.00

For pump 1

Q_1=333.33 gpm H=150

C_1=2.00*8.65*C_{p1}

C_1=8.65*2*276.48 =4783.101$

For pump 2

Q_2=1250 gpm H=150

C_2=2.00*8.65*C_{p2}

C_2=2.00*8.65*669.497 =11582.295$

For pump 3

Q_3=666.67 H=150

C_3=2.00*8.65*C_{p3}

C_3=2.00*8.65*346.008 =5985.95$

From Table the installation factor is 2.00 so that the installed
Price is

Installation cost =potsherd cost*installation factor-----------------------------Eq (4.26)

Cinst (1)= C_1*2 =9566.202$

Cinst (2)= C_2*2 =23164.59$

Cinst (3)= C_3*2 =11971.9$

4.5.1.4 cost of stripers:

C_sFrom Eq (4.16)

C_{pl}FromEq (4.17)

D=4.921

H=8.25

T_b=0.5 in

T_p=0.75 in

N_{tray} =3

W=2813.06

C_P=exp[7.123+0.11478*(Ln2813.06)+0.02488*(Ln2813.06)^2+0.0158*(8.25/4.921)* Ln(0.5/.75)

C_P=14664.31$

F_1=2.1F_2=1.75 F_3=1.59F_4=2

C_{tray} =375.8*exp [0.1739*4.921) from Eq (4.19)

=884.32$

C_{bl}= 204.9*(4.921) ^ (.6332) *8.25^0.8018 from Eq (4.20)

=3051.85$

Cost of striper =2.1*14664.31+1.75*1.59*2*884.32+3051.85

=38768.14$

For 3 striper 3*38768.14

=116,304.42$

4.5.1.5 Cost of condenser:

Assuming (Air cooler)

Cost of Air cooler =24.6*$A^{0.4}$

Where

0.05 <A< 200 K ft^2, price in K$

A=11.25 ft^2

Cost=2.633 $ for 1Btu

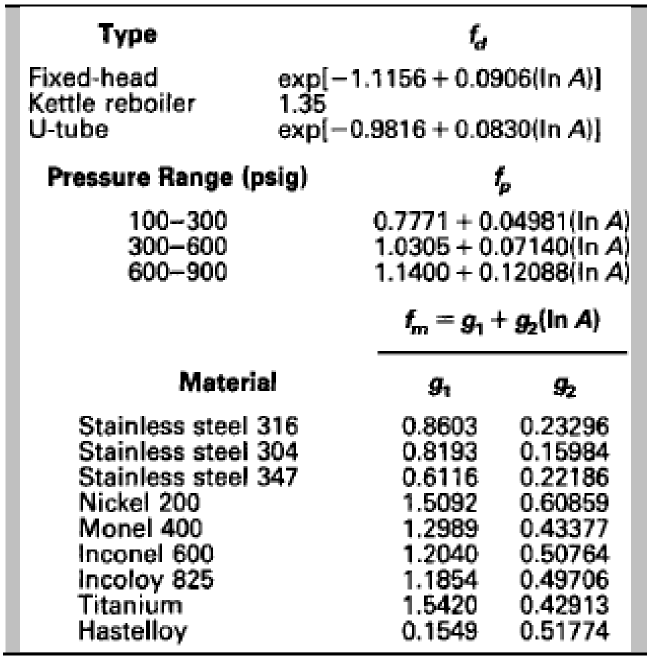
4.5.1.6 Cost of heat exchanger:

C, $= F_d*F_m*F_P ……………………………….Eq(4.27)

C_p=exp [8.821-0.30863*(ln A)+0.0681*(ln A)^2]…………………Eq(4.28)

Where 150 < A <12000 ft^2

Table 4.20 Correction factor of heat exchanger



Type	f_d
Fixed-head	exp[−1.1156 + 0.0906(ln A)]
Kettle reboiler	1.35
U-tube	exp[−0.9816 + 0.0830(ln A)]

Pressure Range (psig)	f_p
100–300	0.7771 + 0.04981(ln A)
300–600	1.0305 + 0.07140(ln A)
600–900	1.1400 + 0.12088(ln A)

	$f_m = g_1 + g_2(\ln A)$	
Material	g_1	g_2
Stainless steel 316	0.8603	0.23296
Stainless steel 304	0.8193	0.15984
Stainless steel 347	0.6116	0.22186
Nickel 200	1.5092	0.60859
Monel 400	1.2989	0.43377
Inconel 600	1.2040	0.50764
Incoloy 825	1.1854	0.49706
Titanium	1.5420	0.42913
Hastelloy	0.1549	0.51774

From data simulation

A=649.3 ft^2

F_d=exp[-0.9816+0.083*(ln A)] (U-tube) ---------------------------------Eq(4.29)

F_d=0.6414

F_p=1.14+0.12088*(ln A)(600—900)psig

F_p=1.923

F_m=g_1+g_2(ln A)………. (7-14)

g_1=0.8603 ,g_2=0.23296

F_m=2.369

C_p=exp[8.821-0.30863*(Ln A)+0.0681*(Ln A)^2

C_p=15966.83$

C=0.6414*1.923*2.369*15966.83

Pusher Cost (C)=46,654.33$

For heat exchangers=4

4*46654.33=186,617.34$

From Table C the installation factor is 1.9 so that the installed

Price is

C_{inst}=1.9*46654.33 =88643.227$

4.5.2 Other costs:

4.5.2.1 Maintenance cost

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

4.5.2.2 power and steam cost

Power	Factor	Cost
Steam 600 psig	1.13	$4.52
Electricity	0.57	$0.04kw/hr
Cooling water	1.0	$0.03/1000gal

Table 4.21 equipment’s costs

equipment costs ,$	Direct-fired heater	Distillation column	Pumps (3-used)	Striper (3-used)	(4)Heat exchangers
Purchased cost	45,553	706,039.54	22,351.35	116,304.4	186,617.34
Installation cost	59,218.9	1,482,683.044	44,702.7	151,195.746	354,573
Total cost	104,771.9	2,188,722.584	67,054.05	267,500.146	541,190.34

Chapter Five

Conclusion & Recommendations

5.1 Conclusion

In this project , anew methodology is presented for facilitating design of crude distillation column , therefore an evaluation of providing the candidate key components and recoveries is supported by simulation program which is Hysis simulator , then trail and error is applied to identify the best combination of key components and based on that design calculation has been done to give 54 number of trays , 2.4m diameter of the column and 27.6m height of the column .

Because there are no two crude oil alike , then many types of sundaes crude oil (6 samples) with different characteristics has been studied to identify the effect of this variation , the simulated crude oil by hysis program present different quantities of atmospheric distillation products and it shown to be sufficiently accurate to predict the quantities of products.

Different proposal has been examined to deal with effect result from the variation of crudes , controlling some parameters such as pumps around flow rate, reflux ratio and multi feed positions is one of the proposals , the other is to change the properties of crude oil before entering to the distillation by upgrading the crude oil according to the API and sulfur content or diluting the heavy crude oil with much lighter crude oil to produce a blend has properties compatible with the properties under which the column has been designed .

5.2 Recommendations

- It is highly recommended to use software has the ability to show the behavior of the crude inside the distillation column such as more advanced Aspen HYSIS versions.
- There are significant impact of using multi feed locations on the distillation column on increasing the naphtha productivity .
- More flexible distillation column is very recommended to compensate the crude oil variation effect on product quantities

Appendix

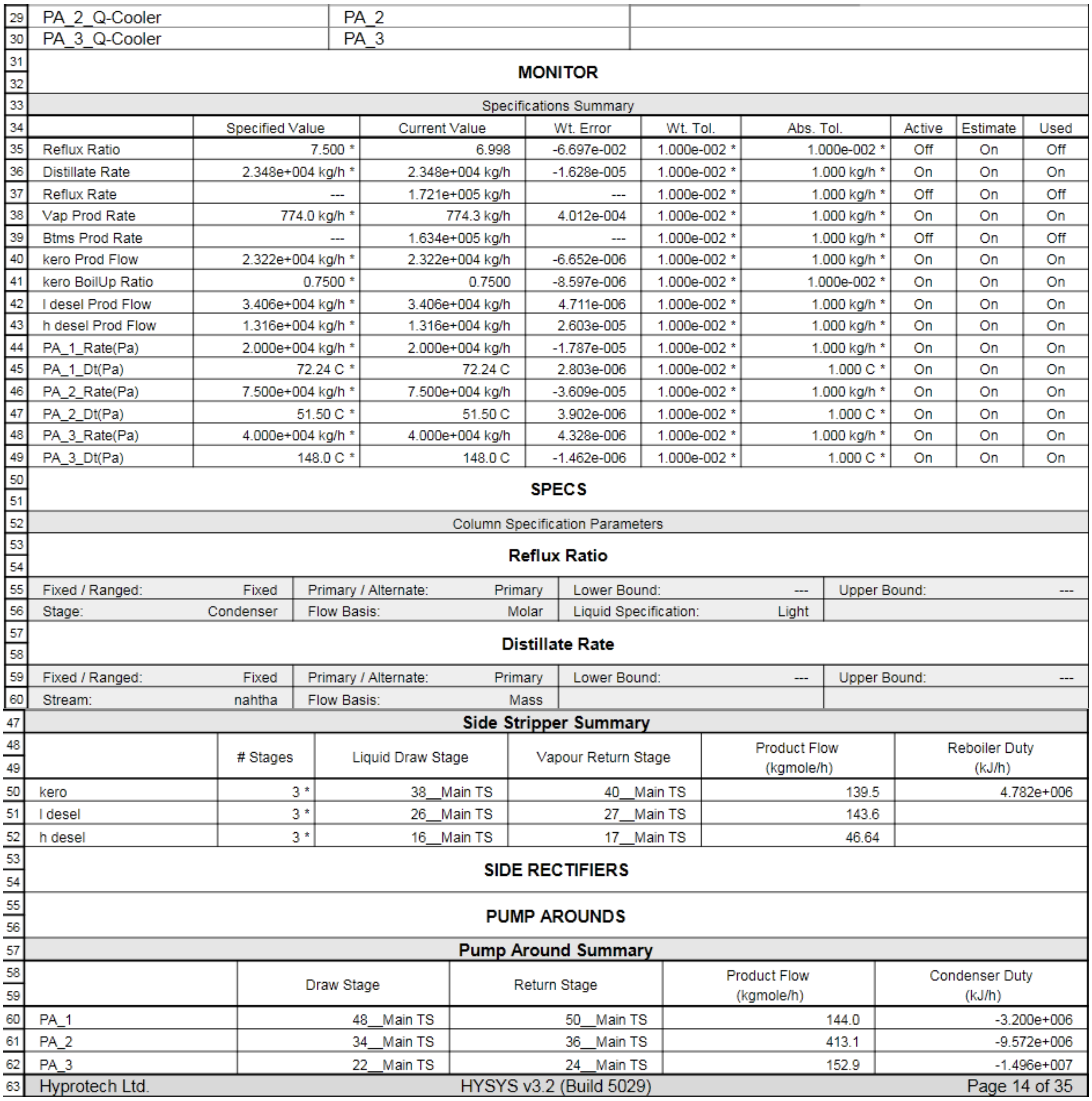
HYSYS reports

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
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000	0.0000



PA_2_Q-Cooler	PA_2	
PA_3_Q-Cooler	PA_3	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Reflux Ratio	7.500 *	6.998	-6.697e-002	1.000e-002 *	1.000e-002 *	Off	On	Off
Distillate Rate	2.348e+004 kg/h *	2.348e+004 kg/h	-1.628e-005	1.000e-002 *	1.000 kg/h *	On	On	On
Reflux Rate	---	1.721e+005 kg/h	---	1.000e-002 *	1.000 kg/h *	Off	On	Off
Vap Prod Rate	774.0 kg/h *	774.3 kg/h	4.012e-004	1.000e-002 *	1.000 kg/h *	On	On	On
Btms Prod Rate	---	1.634e+005 kg/h	---	1.000e-002 *	1.000 kg/h *	Off	On	Off
kero Prod Flow	2.322e+004 kg/h *	2.322e+004 kg/h	-6.652e-006	1.000e-002 *	1.000 kg/h *	On	On	On
kero BoilUp Ratio	0.7500 *	0.7500	-8.597e-006	1.000e-002 *	1.000e-002 *	On	On	On
l desel Prod Flow	3.406e+004 kg/h *	3.406e+004 kg/h	4.711e-006	1.000e-002 *	1.000 kg/h *	On	On	On
h desel Prod Flow	1.316e+004 kg/h *	1.316e+004 kg/h	2.603e-005	1.000e-002 *	1.000 kg/h *	On	On	On
PA_1_Rate(Pa)	2.000e+004 kg/h *	2.000e+004 kg/h	-1.787e-005	1.000e-002 *	1.000 kg/h *	On	On	On
PA_1_Dt(Pa)	72.24 C *	72.24 C	2.803e-006	1.000e-002 *	1.000 C *	On	On	On
PA_2_Rate(Pa)	7.500e+004 kg/h *	7.500e+004 kg/h	-3.609e-005	1.000e-002 *	1.000 kg/h *	On	On	On
PA_2_Dt(Pa)	51.50 C *	51.50 C	3.902e-006	1.000e-002 *	1.000 C *	On	On	On
PA_3_Rate(Pa)	4.000e+004 kg/h *	4.000e+004 kg/h	4.328e-006	1.000e-002 *	1.000 kg/h *	On	On	On
PA_3_Dt(Pa)	148.0 C *	148.0 C	-1.462e-006	1.000e-002 *	1.000 C *	On	On	On

SPECS

Column Specification Parameters

Reflux Ratio

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	Condenser	Flow Basis:	Molar	Liquid Specification:	Light		

Distillate Rate

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	nahtha	Flow Basis:	Mass				

Side Stripper Summary

	# Stages	Liquid Draw Stage	Vapour Return Stage	Product Flow (kgmole/h)	Reboiler Duty (kJ/h)
kero	3 *	38__Main TS	40__Main TS	139.5	4.782e+006
l desel	3 *	26__Main TS	27__Main TS	143.6	
h desel	3 *	16__Main TS	17__Main TS	46.64	

SIDE RECTIFIERS

PUMP AROUNDS

Pump Around Summary

	Draw Stage	Return Stage	Product Flow (kgmole/h)	Condenser Duty (kJ/h)
PA_1	48__Main TS	50__Main TS	144.0	-3.200e+006
PA_2	34__Main TS	36__Main TS	413.1	-9.572e+006
PA_3	22__Main TS	24__Main TS	152.9	-1.496e+007

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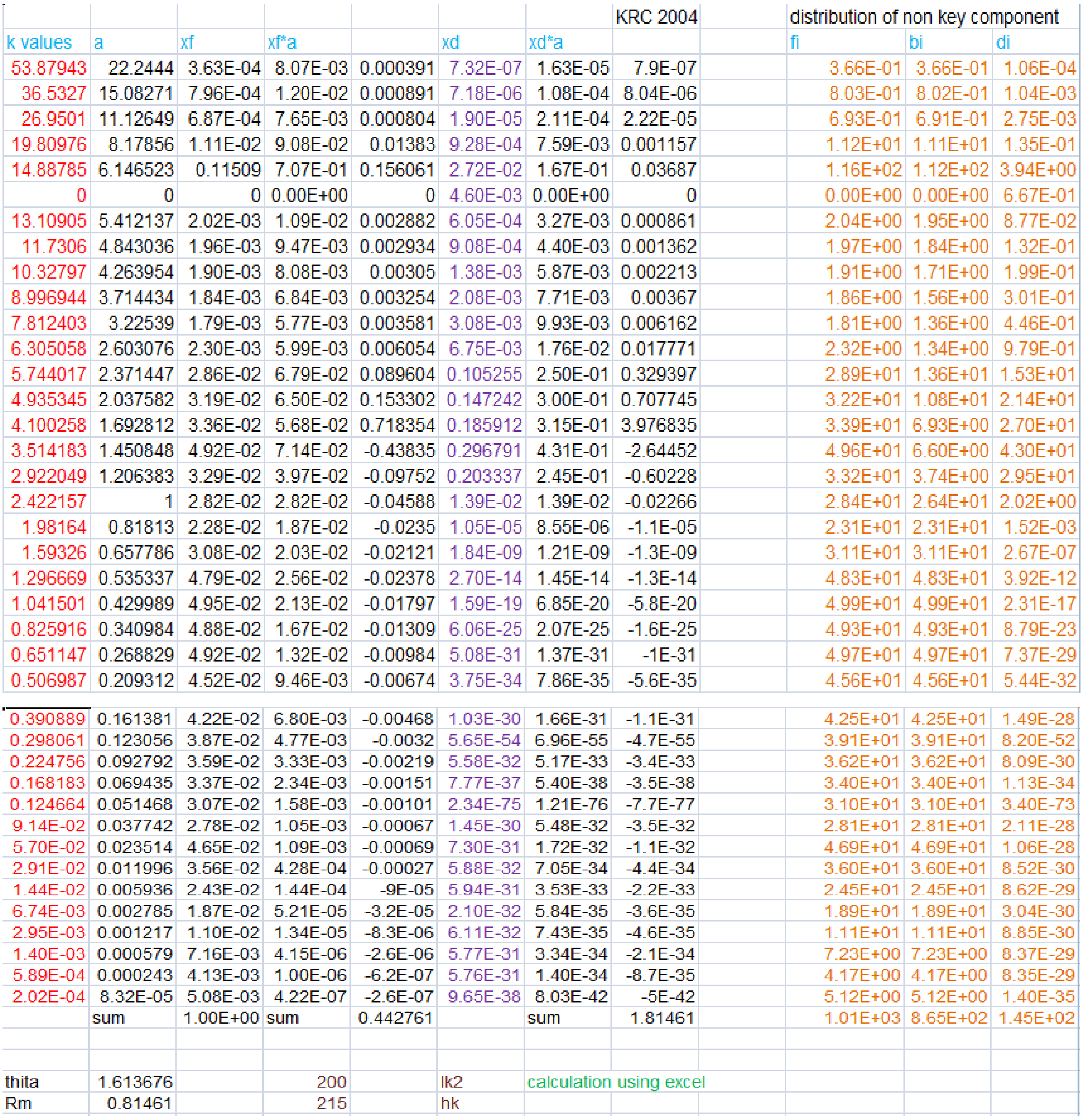
Appendix B

Excel sheet design process



							KRC 2004		distribution of non key component		
k values	a	xf	xf*a		xd	xd*a			fi	bi	di
53.87943	22.2444	3.63E-04	8.07E-03	0.000391	7.32E-07	1.63E-05	7.9E-07		3.66E-01	3.66E-01	1.06E-04
36.5327	15.08271	7.96E-04	1.20E-02	0.000891	7.18E-06	1.08E-04	8.04E-06		8.03E-01	8.02E-01	1.04E-03
26.9501	11.12649	6.87E-04	7.65E-03	0.000804	1.90E-05	2.11E-04	2.22E-05		6.93E-01	6.91E-01	2.75E-03
19.80976	8.17856	1.11E-02	9.08E-02	0.01383	9.28E-04	7.59E-03	0.001157		1.12E+01	1.11E+01	1.35E-01
14.88785	6.146523	0.11509	7.07E-01	0.156061	2.72E-02	1.67E-01	0.03687		1.16E+02	1.12E+02	3.94E+00
0	0	0	0.00E+00	0	4.60E-03	0.00E+00	0		0.00E+00	0.00E+00	6.67E-01
13.10905	5.412137	2.02E-03	1.09E-02	0.002882	6.05E-04	3.27E-03	0.000861		2.04E+00	1.95E+00	8.77E-02
11.7306	4.843036	1.96E-03	9.47E-03	0.002934	9.08E-04	4.40E-03	0.001362		1.97E+00	1.84E+00	1.32E-01
10.32797	4.263954	1.90E-03	8.08E-03	0.00305	1.38E-03	5.87E-03	0.002213		1.91E+00	1.71E+00	1.99E-01
8.996944	3.714434	1.84E-03	6.84E-03	0.003254	2.08E-03	7.71E-03	0.00367		1.86E+00	1.56E+00	3.01E-01
7.812403	3.22539	1.79E-03	5.77E-03	0.003581	3.08E-03	9.93E-03	0.006162		1.81E+00	1.36E+00	4.46E-01
6.305058	2.603076	2.30E-03	5.99E-03	0.006054	6.75E-03	1.76E-02	0.017771		2.32E+00	1.34E+00	9.79E-01
5.744017	2.371447	2.86E-02	6.79E-02	0.089604	0.105255	2.50E-01	0.329397		2.89E+01	1.36E+01	1.53E+01
4.935345	2.037582	3.19E-02	6.50E-02	0.153302	0.147242	3.00E-01	0.707745		3.22E+01	1.08E+01	2.14E+01
4.100258	1.692812	3.36E-02	5.68E-02	0.718354	0.185912	3.15E-01	3.976835		3.39E+01	6.93E+00	2.70E+01
3.514183	1.450848	4.92E-02	7.14E-02	-0.43835	0.296791	4.31E-01	-2.64452		4.96E+01	6.60E+00	4.30E+01
2.922049	1.206383	3.29E-02	3.97E-02	-0.09752	0.203337	2.45E-01	-0.60228		3.32E+01	3.74E+00	2.95E+01
2.422157	1	2.82E-02	2.82E-02	-0.04588	1.39E-02	1.39E-02	-0.02266		2.84E+01	2.64E+01	2.02E+00
1.98164	0.81813	2.28E-02	1.87E-02	-0.0235	1.05E-05	8.55E-06	-1.1E-05		2.31E+01	2.31E+01	1.52E-03
1.59326	0.657786	3.08E-02	2.03E-02	-0.02121	1.84E-09	1.21E-09	-1.3E-09		3.11E+01	3.11E+01	2.67E-07
1.296669	0.535337	4.79E-02	2.56E-02	-0.02378	2.70E-14	1.45E-14	-1.3E-14		4.83E+01	4.83E+01	3.92E-12
1.041501	0.429989	4.95E-02	2.13E-02	-0.01797	1.59E-19	6.85E-20	-5.8E-20		4.99E+01	4.99E+01	2.31E-17
0.825916	0.340984	4.88E-02	1.67E-02	-0.01309	6.06E-25	2.07E-25	-1.6E-25		4.93E+01	4.93E+01	8.79E-23
0.651147	0.268829	4.92E-02	1.32E-02	-0.00984	5.08E-31	1.37E-31	-1E-31		4.97E+01	4.97E+01	7.37E-29
0.506987	0.209312	4.52E-02	9.46E-03	-0.00674	3.75E-34	7.86E-35	-5.6E-35		4.56E+01	4.56E+01	5.44E-32
0.390889	0.161381	4.22E-02	6.80E-03	-0.00468	1.03E-30	1.66E-31	-1.1E-31		4.25E+01	4.25E+01	1.49E-28
0.298061	0.123056	3.87E-02	4.77E-03	-0.0032	5.65E-54	6.96E-55	-4.7E-55		3.91E+01	3.91E+01	8.20E-52
0.224756	0.092792	3.59E-02	3.33E-03	-0.00219	5.58E-32	5.17E-33	-3.4E-33		3.62E+01	3.62E+01	8.09E-30
0.168183	0.069435	3.37E-02	2.34E-03	-0.00151	7.77E-37	5.40E-38	-3.5E-38		3.40E+01	3.40E+01	1.13E-34
0.124664	0.051468	3.07E-02	1.58E-03	-0.00101	2.34E-75	1.21E-76	-7.7E-77		3.10E+01	3.10E+01	3.40E-73
9.14E-02	0.037742	2.78E-02	1.05E-03	-0.00067	1.45E-30	5.48E-32	-3.5E-32		2.81E+01	2.81E+01	2.11E-28
5.70E-02	0.023514	4.65E-02	1.09E-03	-0.00069	7.30E-31	1.72E-32	-1.1E-32		4.69E+01	4.69E+01	1.06E-28
2.91E-02	0.011996	3.56E-02	4.28E-04	-0.00027	5.88E-32	7.05E-34	-4.4E-34		3.60E+01	3.60E+01	8.52E-30
1.44E-02	0.005936	2.43E-02	1.44E-04	-9E-05	5.94E-31	3.53E-33	-2.2E-33		2.45E+01	2.45E+01	8.62E-29
6.74E-03	0.002785	1.87E-02	5.21E-05	-3.2E-05	2.10E-32	5.84E-35	-3.6E-35		1.89E+01	1.89E+01	3.04E-30
2.95E-03	0.001217	1.10E-02	1.34E-05	-8.3E-06	6.11E-32	7.43E-35	-4.6E-35		1.11E+01	1.11E+01	8.85E-30
1.40E-03	0.000579	7.16E-03	4.15E-06	-2.6E-06	5.77E-31	3.34E-34	-2.1E-34		7.23E+00	7.23E+00	8.37E-29
5.89E-04	0.000243	4.13E-03	1.00E-06	-6.2E-07	5.76E-31	1.40E-34	-8.7E-35		4.17E+00	4.17E+00	8.35E-29
2.02E-04	8.32E-05	5.08E-03	4.22E-07	-2.6E-07	9.65E-38	8.03E-42	-5E-42		5.12E+00	5.12E+00	1.40E-35
	sum	1.00E+00	sum	0.442761		sum	1.81461		1.01E+03	8.65E+02	1.45E+02
thita	1.613676		200		lk2	calculation using excel					
Rm	0.81461		215		hk						

Appendix C

COST ESTIMATION

Table C.1Multipliers 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

Appendix D

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

	DRAW STAGE	RETURNE STAGE	DRAW RATE(t/h)	$T_{1\ (C)}$	$T_{2\ (C)}$

Heavy diesel	Steam	16	17

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 Distillation column operating conditions

TOP STAGE	125.9ºC	0.053 Mpa
BOTTOM STAGE	360.7ºC	0.072 Mpa

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

.

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