بِسۡمِ ٱللَّهِ ٱلرَّحۡمَٰنِ ٱلرَّحِيمِ

Sudan University of Science and Technology

College of Petroleum Engineering and Technology

Department of Transportation and Refining Engineering

SIMULATION OF VACUUM DISTILLATION UNIT (VDU) AND IMPROVING ITS PRODUCTIVITY USING ASPEN HYSYS

محاكاة وحدة التقطير الفراغي وتحسين إنتاجيتها باستخدام برنامج الهايسيس

Dissertation Submitted in Partial Fulfillment of the Requirement for the Bachelor of Engineering (Horns) Degree in Transportation and Refining Engineering

Prepared by:

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Date: / /2015

الإستهلال

قال تعالى:

" وَمَا تَوْفِيقِي إِلَّا بِاللَّهِ ۚ عَلَيْهِ تَوَكَّلْتُ وَإِلَيْهِ أُنِيبُ "

سورة هود الاية (88)

Dedication

We dedicate this project to our parents for the love and support they have provided throughout our entire life, they have been there for every decision we have made and help our dreams become reality, to our friends and families for their help and encouragement. We also dedicate this project to our dear friends ***Ms. MuazTajaldeen*** and ***Ms. Mohamed Abuelgasem***

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Abstract

The project studies the design of vacuum distillation unit (VDU) to process the atmospheric residue produced from Khartoum Refinery. It is simulated on Aspen HYSYS v8 Program. The project includes material and energy balance, designing the vacuum distillation unit (VDU), estimating cost of vacuum distillation unit, corrosion protection of the vacuum distillation and improving the production of (VDU).Cost estimation with the economic summary from ASPEN hysys allows us to assess the total cost of the unit. According to this study, it is found that it is possible to increase the production of heavy vacuum gas oil (HVGO) and light vacuum gas oil (LVGO) and this is a main objective we work for it.

Keywords: Vacuum distillation unit, production of gas oil, cost estimation, ASPENHysys, process design.

المستخلص

يدرس هذا المشروع تصميم و محاكاة وحده التقطير الفراغي لمعالجة المتبقي الجوي الناتج من مصفاة الخرطوم, اعتمد المشروع على برنامج هايسسAspen Hysys V8 و موازنة المادة و الطاقة و تصميم وحده التقطير الفراغي و تقدير التكلفة و تحسين الانتاجية وطرق حماية المعدات من التاكل. وفقا لهذه الدراسة وجد انه من الممكن تحويل نسبة كبيره من المتبقي الجوي الي و أيضا تحسين و تطوير هذا الانتاج عن طريق التحكم في ظروف التشغيل يؤدي لزياد ملحوظ في زيت الخام الثقيل (HVGO) وزيت الخام الخفيف (LVGO).

Table of Contents:

Chapter 1:Introduction

CHAPTER 2: Literature Review

Chapter 3: Methodology

Chapter 4: Results & Discussion

Chapter 5: Conclusion &Recommendation

List of Figures:

List of Tables:

Chapter1

INTRODUCTION

1.1. Introduction:

Numerous methods have been developed to selectively remove water and other volatile contaminants from hydraulic and lubricating fluids. These methods include absorbent filter media and regenerable adsorbent packings and In many cases, it is not economical or practical to use disposable media, and as a result, continuous scrubbing processes have been developed. The most common scrubbers are derivatives of vacuum distillation processes used in refineries.

In this project we will design the vacuum distillation unit for process the atmospheric residue form Khartoum refinery (KRF). ASPEN Hysys software will be used in simulation of vacuum distillation column gives us accurate Information of how processes take place.

Material and energy balance will concentrate on the quantity of energy required by units and mass flow of matter during process. In the cost estimation we will estimate total cost of operation.

The main reason to do this project is to try to improve the production of this column after designing the vacuum distillation column. We will concentrate on improving gas oil production because it is one of the most valuable products.

1.2 Aspen Hysys:

Aspen HYSYS Refining contains a database, the petroleum assay, that you can use to store and calculate the physical and petroleum properties of the crude oil stream. The petroleum assay is a vector that stores physical properties and assay properties for a specific component list. Physical properties include all properties used in a typical HYSYS simulation case. Assay properties comprise refinery related properties as cloud point, octane numbers, flash point, freezepoint, sulphur content, PONA distribution, GC data and etc. A component list typically consists of library components (for instance, methane to n-pentane) and pseudo-components(hypothetical components). Aspen HYSYS Refining is based on a flexible structure so that no pre-defined list of pseudo component sis required. Moreover, existing lists of pseudo components created by the HYSYS Oil Environment can be used in Aspen HYSYS Refining. Each component stores a value of a physical and assay property. The assay properties are usually imported from an assay management system, as for instance, Crude Manager(TM)-Aspen HYSYS Refining Link from Spiral Software Ltd. At the Simulation Environment, each stream may have its own petroleum assay, that is, the physical and assay properties of components on one stream may differ from other streams. Bulk values for assay properties are calculated using specific lumping rules. When process streams are mixed together on any HYSYS or Aspen HYSYS Refining operation, a new petroleum assay is created and special blending rules are employed to re-calculate the physical and assay properties.

HYSYS Refining is based upon a flexible structure that allows auser to characterize this petroleum assay using the least available data and using rigorous laboratory data. The user does not need to have data in the RefSYS format mentioned above; RefSYS can take data in different formats and transmute it into an internal petroleum assay format.

1.3 Vacuum distillation column:

Atmospheric residue (AR) from atmospheric distillation tower contains several valuable cuts which should be recovered. AR cannot be fractionated at atmospheric tower as fractionation of this cut needs excessive temperature where, cracking or decomposition of crude starts resulting in severe coke deposition. Hence, AR is recovered as a bottom product from the tower and distilled under sub-atmospheric pressure.

Crude oil can be categorized as lube-bearing crude and non-lube bearing crude. Non-lube bearing crude cannot produce lubricating oil cut in vacuum distillation as this range of hydrocarbons are not present in non-lube bearing crude. The unit for processing of non-lube bearing crude and lube-bearing crude are known as fuel-type and lube-type vacuum distillation column respectively. The former produces vacuum gas oil (VGO) and later produces lubricating oil as the main distillate product. AR is introduced into the vacuum distillation column after heat exchanging with distillation products, vacuum residue and pump-around reflux streams and finally heated in a furnace at required temperature. Vacuum distillation furnace may be classified into two types, wet and dry. In wet type, steam is injected into the furnace coils and that helps to lower the partial pressure of feed as well as steam carries the feed vapours through the furnace tube more rapidly. In dry type, steam injection is not done in the furnace. Steam injection lowers the steam consumption in the vacuum ejector systems. The choice of the type depends on the overall economy of the refinery.

1.4 Objectives:

-This project aims to operate atmospheric distillation residue on a vacuum distillation to produce gas oil.

-Detailed design for vacuum distillation column.

-Improve the production of light vacuum gas oil (LVGO) and heavy gas oil (HVGO).

1.5 Scope of this study:

The scope of this project is to give detailed study for design of vacuum distillation unit. The project will cover the following:

1- ASPEN Hysys software.

2- Material balance.

3- Energy balance.

4- Process design.

5- Cost estimation.

All calculation are performed by Excel worksheet.

CHAPTER 2

Literature Review

2.1 Oil refinery

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

Raw or unprocessed crude oil is not generally useful in industrial applications, although "light, sweet" (low viscosity, low sulfur) crude oil has been used directly as a burner fuel to produce steam for the propulsion of seagoing vessels. The lighter elements, however, form explosive vapors in the fuel tanks and are therefore hazardous, especially in warships. Instead, the hundreds of different hydrocarbon molecules in crude oil are separated in a refinery into components which can be used as fuels, lubricants, and as feedstock's in petrochemical processes that manufacture such products as plastics, detergents, solvents, elastomers and fibers such as nylon and polyesters.

2.2 Refining process:

Most refineries, regardless of complexity, perform a few basic steps in the refining process: DISTILLATION, CRACKING, TREATING and REFORMING. These processes occur in our main operating areas – Crude/Aromatics, Cracking I, RDS/Coker, Cracking II, and at the Sulfur Recovery Unit.

2.2.1 Distillation

Modern distillation involves pumping oil through pipes in hot furnaces and separating light hydrocarbon molecules from heavy ones in downstream distillation towers – the tall, narrow columns that give refineries their distinctive skylines.

The Pascagoula Refinery's refining process begins when crude oil is distilled in two large Crude Units that have three distillation columns, one that operates at near atmospheric pressure, and two others that operate at less than atmospheric pressure, i.e., a vacuum.

During this process, the lightest materials, like propane and butane, vaporize and rise to the top of the first atmospheric column. Medium weight materials, including gasoline, jet and diesel fuels, condense in the middle. Heavy materials, called gas oils, condense in the lower portion of the atmospheric column. The heaviest tar-like material, called residuum, is referred to as the "bottom of the barrel" because it never really rises.

This distillation process is repeated in many other plants as the oil is further refined to make various products.

In some cases, distillation columns are operated at less than atmospheric pressure (vacuum) to lower the temperature at which a hydrocarbon mixture boils. This "vacuum distillation" (VDU) reduces the chance of thermal decomposition (cracking) due to overheating the mixture.

Using the most up-to-date computer control systems, refinery operators precisely control the temperatures in the distillation columns which are designed with pipes to withdraw the various types of products where they condense. Products from the top, middle and bottom of the column travel through these pipes to different plants for further refining.

2.2.2 Cracking

Convert middle distillate, gas oil and residuum into primarily gasoline, jet and diesel fuels by using a series of processing plants that literally "crack" large, heavy molecules into smaller, lighter ones.

Heat and catalysts are used to convert the heavier oils to lighter products using three "cracking" methods: fluid catalytic cracking (FCC), hydrocracking (Isomax), and coking (or thermal-cracking).

The Fluid Catalytic Cracker (FCC) uses high temperature and catalyst to crack heavy gas oil mostly into gasoline. Hydrocracking uses catalysts to react gas oil and hydrogen under high pressure and high temperature to make both jet fuel and gasoline.

We blend most of the products from the FCC and the Isomaxes directly into transportation fuels, i.e., gasoline, diesel and jet fuel. We burn the lightest molecules as fuel for the refinery's furnaces, thus conserving natural gas and minimizing waste.

In the Delayed Coking Unit (Coker) , low-value residuum is converted (using the coking, or thermal-cracking process) to high-value light products, producing petroleum coke as a by-product. The large residuum molecules are cracked into smaller molecules when the residuum is held in a coke drum at a high temperature for a period of time. Only solid coke remains and must be drilled from the coke drums.

2.2.3 Combining

While the cracking processes break most of the gas oil into gasoline and jet fuel, they also break off some pieces that are lighter than gasoline. This process takes the small molecules and recombines them in the presence of sulfuric acid catalyst to convert them into high octane gasoline.

2.2.4 Treating (Removing Impurities)

The products from the Crude Units and the feeds to other units contain some natural impurities, such as sulfur and nitrogen. Using a process called hydrotreating (a milder version of hydrocracking), these impurities are removed to reduce air pollution when the fuels are used

In the RDS Unit's six 1,000-ton reactors, sulfur and nitrogen are removed from FCC feed stream. The sulfur is converted to hydrogen sulfide and sent to the Sulfur Unit where it is converted into elemental sulfur. Nitrogen is transformed into ammonia which is removed from the process by water-washing. Later, the water is treated to recover the ammonia as a pure product for use in the production of fertilizer.

The RDS's Unit main product, low sulfur vacuum gas oil, is fed to the FCC (fluid catalytic cracker) Unit which then cracks it into high value products such as gasoline and diesel.

2.2.5 Reforming

Octane rating is a key measurement of how well a gasoline performs in an automobile engine. Much of the gasoline that comes from the Crude Units or from the Cracking Units does not have enough octane to burn well in cars.

The gasoline process streams in the refinery that have a fairly low octane rating are sent to a Reforming Unit where their octane levels are boosted. These reforming units employ precious-metal catalysts - platinum and rhenium – and thereby get the name "rheniformers." In the reforming process, hydrocarbon molecules are "reformed" into high octane gasoline components. For example, methyl cyclohexane is reformed into toluene.

The reforming process actually removes hydrogen from low-octane gasoline. The hydrogen is used throughout the refinery in various cracking (hydrocracking) and treating (hydrotreating) units.

2.2.6 Blending

A final and critical step is the blending of our products. Gasoline, for example, is blended from treated components made in several processing units. Blending and Shipping Area operators precisely combine these to ensure that the blend has the right octane level, vapor pressure rating and other important specifications. All products are blended in a similar fashion.

2.2.7 Quality Control

In the refinery's modernly-equipped Laboratory, chemists and technicians conduct quality assurance tests on all finished products, including checking gasoline for proper octane rating.

2.3 Process Units in KRC:

- Crude Distillation Unit (CDU).
- Residue Catalytic Cracking Unit (RFCC).
- Semi-regeneration Catalytic Reforming Unit (SCR).
- Diesel Hydro Treating Unit (DHT).
- Delayed Coking Unit (DCU).
- Gasoline/ Diesel Hydro Treating Unit (GDHT).
- Continuous Catalytic Reforming Unit (CCR)

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1. Crude Distillation Unit (CDU)

To produce naphtha (which it the feed (SCR)), jet fuel, diesel and residue the feed of (RFCC) unit.

2. Residue Catalytic Cracking Unit (RFCC).

It designed to process the residue from (CDU) and consist of five parts:-

The reactor-regenerator, the fractionation, the absorber -diabsorber, sweating system for gasoline, LPG and dry gas and compressors. The products of (RFCC) are dry gas, LPG, gasoline, gas oil and slurry.

3. Semi-regeneration Catalytic Reforming Unit (SCR)

It designed to process the naphtha from (CDU) to produce the refined gasoline and at the same time to supply the hydrogen to (DHT). This unit had been shut down after (CCR) unit run normal and now it is standby unit.

4. Diesel Hydro Treating Unit (DHT)

It designed to process the Gas oil from (RFCC) and to produce refined diesel.

5. Delayed Coking Unit (DCU)

It designed to process Fulla Crude Oil and its products include fuel gas, LPG, Coker gasoline , Coker diesel, the heavy coked gas oil and the petroleum Coke.

6. Gasoline/ Diesel Hydro Treating Unit (GDHT)

ton it designed to treat the Coker gasoline and the Coker diesel from (DCU) unit to produce LPG, the refined diesel and treated Coker gasoline.

7. Continuous Catalytic Reforming Unit (CCR).

it designed to treat the mixture of the treated Coker gasoline and (CDU) unit naphtha and produce LPG, gasoline and hydrogen which it is send to (GDHT) and (DHT).

2.4 Petroleum Products and Characterization:

2.4.1 Gas

Gas from petroleum is produced and classified under different names

2.4.1.1 Natural (kuff) gas: Produced as free gas from the reservoir. It contributes 85-98% methane accompanied by ethane, condensate, inert ($N2/CO2$) and sulfur.

2.4.1.2 Associated gas: Found as separate gas cap above oil level inside the reservoir. The pressure of this gas is the source energy of the primary oil recovery from the well. It is mainly methane with proportions of C2/C3/C4/C5 recovered as NGL (natural gas liquid)

2.4.1.3 Dissolved gas: Found dissolved in oil due to high pressure. It contributes up to 600 $SCFt^3$/bbl. For piping and storage of crude oil, this gas should be separated at early stages of crude processing, namely at GOSP and CPF.

2.4.1.4 Refinery off gas: developed from conversion, stabilization and topping processes like reforming and cracking. It constitutes:

- Dry gas (C1/C2) which is utilized as refinery fuel gas.
- Wet gas (C3/C4/C5) utilized as LPG sales and gasoline blending
- Impurities like CO2, N2, mercaptans, H2S and water vapor.

Each of the above classes may be utilized as petrochemical feedstock. Sulfur may be recovered from gas when it is in considerable amount.

2.4.2 LPG (liquefied petroleum gas)

It is produced from different refinery processes e.g. cracking, reforming. It is a mixture of propane and butane in some different gradients. It widely used as domestic fuel that is supplied in 12-15kg cylinders, beside petrochemical feed stocks. It is stored in vapor/liquid mixture.

Important characteristics of LPG

- Vapor pressure at 65C = 10-26kg/cm2
- Reid Vapor Pressure RVP = 6.7 kg/cm2
- 95% evaporation temperature = -2C
- Sulfur % wt = 0.02
- H2S = Nil
- Moisture content = Nil

LPG Composition

component	value
Ethane	Traces
Propane	24.7%
i-butane	36.7%
butane	36.5%
pentane	nil

2.4.3 Gasoline

It is volatile petroleum fraction used as motor fuel (motor spirit) with boiling range from 37 0C to 180 0C. It is produce as:

- Straight run gasoline (SRG, C5-93 0C) from distillation. It is of low octane number. It is used as fuel blend.
- Reformate gasoline from reforming process
- Coker gasoline from thermal cracking e.g. DCU
- Alkylated gasoline
- Polymer gasoline
- Cracked gasoline from catalytic / hydro-cracking.

Gasoline is characterized by ASTM distillation range, RVP, octane number, gum and sulfur contents.

2.4.4 Naphtha

Volatile cut with boiling range of the motor spirit. It is of low octane number. It is mainly used in reforming process to produce high octane gasoline. Naphtha may be used as paint solvent.

2.4.5 Kerosene

Boiling range =150-250 C. It is used as heating oil and in jet production. Kerosene is characterized by:

Flash point: usually above 42 C

Smoke point: it is defined as the max height of the flame in mm at which the given oil will burn without giving smoke.

Volatility: 10% evaporation temp indicates flash point and ease of ignition

50% evaporation temp indicates viscosity

60% evaporation temp indicates steady performance

Sulfur content: max of 0.13%wt is accepted

Burning Quality: good oil of 650 cc must burn for at least 120 hrs

Aniline point: indicates aromatic content

2.4.6 Diesels

These are gas oil fractions in the boiling range 250-320 C

Characterized by:

Aniline point: it is defined as the min temperature at which equal volumes of anhydrous Aniline and oil mix together. It is an indication for aromatics content in oil. Aromatics in fuel cause abnormal ignition Diesel Index = 45 to 55 normal

Flash Point: it is of no real significance on fuel performance, but required for safe storage and handling. Usually 50-55 0C

Calorific Value: 41.83 kj /gm

Viscosity: ease of start and low temp operability

2.4.7 Lub Oil

These are fractions left after light components from crude distillation. They are of boiling range above 350 0C. They are obtained from vacuum distillation of atmospheric distillation residue.

General Molecular Structure

Naphthenic or aromatic rings arranged in as many as six groups with Paraffinic side chain.

Lube oils are characterized by: flash point, pour point, oxidation stability, viscosity and carbon residue.

2.4.8 Bitumen

It is residual product from crude distillation (bottom of the barrel). It is solid at room temperature. Bitumen is blown with air to produce asphalt utilized in highway construction, water proofing and coating works. Bitumen is characterized by: softening point, penetration index and ductility. Ductility is the capacity to elongation and stretching.

2.4.9 Coke

Solid coke from crude is produced by thermal cracking process. Typical application is Fula blend cracking in KRC-Coker of 14%wt yield on crude basis.

2.5 Crude assay:

Crude oils and petroleum fractions are the most important feed stocks for refining processes. To properly simulate the refining processes, we must have good understanding of the compositional information and thermo physical properties of crude oils and petroleum fractions.

However, the complexity of molecular composition of crude oils and petroleum fractions makes it hardly possible to identity individual molecules. Instead, modern refiners use assay to characterize crude oil sand petroleum fractions.*(A.-F. Chang, K. 2012)*

It is extracted from the earth, formed naturally from the fossil of animals and plants. The viscosity and relative weight of crude oil varies and it can exist in either liquid or solid state.

Color: Light brown to dark brown

Sp.gr: 0.81—0.985

Boiling range: 25 – 400oC

Hydrocarbons C1- C70 (4000 compounds)

Metals: V, Fe, Ni S-(H2S,Thiols (mercaptans), sulfides, di sulfides, poly sulfides and thiophenes).Cause corrosion of equipment's, bad odour in products, catalyst poisoning, Air pollution –Indols, pyridines and quinolenes (Difficult to remove).

- Oxygen compounds: present as naphthenic acids and phenols
- Are corrosive in nature and cause odour.
- Metal: act as catalyst poisons

Crude composition

C: 84-87%, H: 11-14%,S: 0-5%,N: 0-1%,O: 0-2%.

Crude assay is a detailed report which describes the properties of the whole crude, as well as the major fractions into which a crudes distilled at the refinery -gasoline, naphtha, kerosene, jet fuel, middle distillates, gas oils and reside. Typically, the data contained in a crude assay includes yields generated from the physical distillation & Distillate/reside properties. (*U.VenkataRamana," 25th–28 AUG 2010)*

A typical crude assay includes two types of information for an oil sample:

(1) bulk properties; and (2) fractional properties. For design and modeling purposes, it is always the best practice to have process data obtained in the same period as assay data, since the properties and composition of crude change over time as it is produced from a given well

2.5.1 Bulk properties:

Bulk properties include specific gravity, sulfur content, nitrogen content, metal (Ni, V, Fe etc.) content, asphaltene content, C/H ratio, pour point, flash point, freeze point, smoke point, aniline point, cloud point, viscosity, carbon residue, light hydrocarbon yields (C1–C4), acid number, refractive index and boiling point curve. We generally use the API (American Petroleum Institute) gravity to specify the specific gravity (SG) of the crude oil as API = (141.5/SG) – 131.5. SG is the specific gravity defined as the ratio of the density of the crude oil to the density of water both at 15.6 °C (60 °F). The API gravity varies from less than 10 for very heavy crudes, to between 10 and 30 for heavy crudes, to between 30 and 40 for medium crudes, and to above 40 for light crudes.

The sulfur content is expressed as a percentage of sulfur by weight, and varies from less than 0.1% to greater than 5%. Crude oils with less than 1 wt.% sulfur are called low-sulfur or sweet crude, and those with more than 1 wt.% sulfur are called high-sulfur or sour crude. Sulfur-containing constituents of the crude oil include simple mercaptans (also known as thiols), sulfides, and polycyclic sulfides. Mercaptan sulfur is simply an alkyl chain (R–) with –SH group attached to it at the end. The simplest form of R–SH is methyl mercaptan, CH3SH.

The pour point is a measure of how easy or difficult to pump the crude oil, especially in cold weather. Specifically, the pour point is the lowest temperature at which a crude oil will flow or pour when it is chilled without disturbance at a controlled rate. The pour point of the whole crude or oil fractions boiling above 232 °C (450 °F) is determined by the standard test ASTM D97.

The flash point of a liquid hydrocarbon or an oil fraction indicates its fire and explosion potential, and it is the lowest temperature at which sufficient vapor is produced above the liquid to form a mixture with air that a spontaneous ignition can occur if a spark is present. One of the standard ASTM test methods for the flash point is D3278.

The freeze point is the temperature at which the hydrocarbon liquid solidifies at atmospheric pressure. It’s an important property for kerosene and jet fuels, because of the very low temperatures encountered at high altitudes in jet planes. One of the standard test methods for the freeze point is ASTM D4790.

The smoke point refers to the height of a smokeless flame of fuel in millimeters beyond which smoking takes places. It reflects the burning quality of kerosene and jet fuels, and is determined by the standard testing method ASTM D1322.

The aniline point represents the minimum temperature for complete miscibility of equal volumes of aniline and petroleum oil. It’s an important property of diesel fuels, and is measured by ASTM D611.

The cloud point refers to the temperature at which solidifable components (waxes) present in the oil sample begin to crystallize or separate from solution under a method of prescribed chilling. It’s an important specification of middle distillate fuels, as determined by ASTM D2500.

The Conrad son carbon residue (CCR) results from ASTM test D189. It measures the coke-forming tendencies of oil. It is determined by destructive distillation of a sample to elemental carbon (coke residue), in the absence of air, expressed as the weight percentage of the original sample. A related measure of the carbon residue is called Rams bottom carbon residue. A crude oil with a high CCR has allow value as a refinery feedstock.

The acid number results from ASTM test method D3339-11 that determines the organic acidity of a refinery stream.

The refractive index represents the ratio of the velocity of light in a vacuum to that in the oil. It is determined by ASTM D1218.

The gross heat of combustion or high heating value (HHV) is the amount of heat produced by the complete combustion of a unit quantity of fuel. We obtain the gross heat of combustion by cooling down all products of the combustion to the temperature before the combustion, and by condensing all the water vapor formed during combustion.

The net heat of combustion or lower heating value (LHV) is obtained by subtracting the latent heat of vaporization of the water vapor formed by the combustion from the gross heat of combustion or higher heating value.

The true boiling point (TBP) distillation of a crude oil or petroleum fractions results from using the U. S. Bureau of Mines Hempel method and the ASTM D-285 test procedure. Neither of these methods specifies the number of theoretical stages or the molar reflux ratio used in the distillation. Consequently, there is a trend toward applying a 15 : 5 distillation according to ASTM D2892, instead of the TBP. The 15 : 5 distillation uses 15 theoretical stages and a molar reflux ratio of 5. A key result from a distillation test is the boiling point curve, that is, the boiling point of the oil fraction versus the fraction of oil vaporized. The initial boiling point (IBP) is defined as the temperature at which the first drop of liquid leaves the condenser tube of the distillation apparatus. The final boiling point or the end point (EP) is the highest temperature recorded in the test.

The ASTM D86 distillation of an oil fraction takes place at laboratory room temperature and pressure. Note that the D86 distillation will end below an approximate temperature of 650 °F (344 °C), at which petroleum oils begin to crack at one atmospheric pressure.

The ASTM D1160 distillation of an oil fraction is applicable to high-boiling oil samples (e.g. heavy heating oil, cracker gas oil feed, residual oil, etc.) for which there is significant cracking at atmospheric pressures. The sample is distilled at a reduced pressure, typically at 10 mmHg, to inhibit cracking. In fact, at 10 mmHg, we can distill an oil fraction up to temperatures of 950 to 1000 °F (510 to 538 °C),as reported on a 760-mmHg basis. The reduced pressure used for D1160 distillation produces a separation of components that is more ideal than that for D86 distillation. *(A.-F. Chang, 2012)*

2.5.2 Fractional Properties

Refineries require fractional properties of the oil sample that reflects the property and composition for specific boiling point range to properly refine it into different end products such as gasoline, diesel and raw materials for chemical process. Fractional properties usually contains paraffins, naphthenes and aromatics (PNA) contents, sulfur content, nitrogen content for each boiling-point range, octane number for gasoline, freezing point, cetane index and smoke point for kerosene and diesel fuels.

The octane number is a measure of the knocking characteristics of a fuel in a laboratory gasoline engine according to ASTM D2700 [1]. We determine the octane number of a fuel by measuring its knocking value compared to the knocking of a mixture of n-heptane and isooctane or 2-2-4-trimethylpentane (224TMP). By definition, we assign an octane number of 0 to pure heptane and of 100 to 224TMP. Therefore, a mixture of 30% heptanes and 70% isooctane has an octane number of 70.

There are two specific octane numbers in use. The motor octane number (MON) reflects the engine performance at highway conditions with high speeds (900 rpm), while the research octane number (RON) corresponds to the low-speed city driving (600 rpm). RON is typically higher than MON because of engine test efficiencies. The posted octane number is an average of MON and RON.

The cetane number measures the ease for self-ignition of a diesel fuel sample and is essentially an opposite of the octane number. It represents the percentage of pure cetane (n-hexadecane) in a blend of cetane and alpha methyl-naphthalene that matches the ignition quality of a diesel fuel sample. This quality is important for middle distillate fuels.

The cetane index is a number calculated from the average boiling point and gravity of a petroleum fraction in the diesel fuel boiling range, which estimates the cetane number of the fraction according to ASTM D976. *(A.-F. Chang, 2012)*

2.6 Previous work:

Distillation was developed into its modern form with the invention of the alembic by Islamic alchemist Jabir ibnHayyan in around 800 AD. The distilling of petroleum products from crude oil to some extent or other has long been practiced. Certainly the ancient Egyptians, Greeks, and Romans had some form of extracting a flammable oil from, probably, weathered crude oil seepage. It wasn’t though until the turn of the nineteenth and twentieth century that crude oil well drilling was first discovered and commercialized. It will be of interest to outline briefly an important development that occurred in this process during the early 1960s. Originally vacuum units followed closely on design to the atmospheric unit except of course it operated under a vacuum condition.

2.6.1 Deep Cut Vacuum Tower Incentives for Various Crudes:

This study compares deep cut incentives for typical light and heavy crudes, Brent and Arabian Heavy. The resultant yields structures and incentives are compared for the two crudes, and economic calculations presented for various charge rates. This technique is valuable in determining project potential and equipment requirements. Deep Cut concerns are also reviewed so that a structure for benefit and risk analysis can be developed. *(Donald F. Schneider, March 1997)*

2.6.2 Feed Characterization and Deepcut Vacuum Columns: Simulation and Design:

Impact of High Temperature Simulated Distillation: The higher vacuum gas oil yield associated with deepcut improves refinery margins. The amount of additional Gas oil is a function of the processing conditions and the quantity of recoverable gas oil in the feed. he vacuum column feed distillation can be improved by using gas chromatographic ASTMD2887 and high-temperature simulated distillations (HTSD). *(S.W. Golden, , April 199)*

2.6.3 Improving crude vacuum unit performance:

This article focuses on sharing data specifically from unconventional heavy crude vacuum units operating in wet mode (stripping and velocity steam) to provide some insight into using a proven vacuum distillation tower simulation topology with heavy oil/bitumen upgrade feeds. *(Darius Remesat Q3 2008,)*

2.7 Process description:

The atmospheric residue coming from atmospheric distillation unit is sent directly to vacuum distillation unit. The residue is sometimes stored at range of 150 °C to guarantee its viscosity. It's heat exchanged against hot product and pumparound streams before being vaporized in the distillation unit heater. Afterward it is heated in a furnace at a maximum temperature of some 380 to 415 °C at the column inlet and feed into vacuum distillation unit.

The distillated products of atmospheric distillation unit (ADU) are limited to the boiling fractions under 350 °C such as gasoline and diesel because petroleum fractions tend to thermally degrade in high temperatures. To recover additional distillates and gas oils, the refinery uses vacuum distillation unit (VDU) following the ADU. The reduced operating pressure of VDU allows recovery of heavy boiling fraction above 560 °C from the atmospheric residue.

There are two major types of VDU operations in a modern refinery – *feedstock preparation and lubricant production*. Feedstock preparation is the most common operation that recovers gas oil from the atmospheric residue as a feed to the downstream conversion units (e.g. FCC and hydrocracking units), which converts the gas oil into more valuable liquid products such as gasoline and diesel. Lubricant production is designed to extract petroleum fractions from the atmospheric residue to produce luboil with desirable viscosity and other related properties.

Thereafter the distillate vapors are condensed in the tower by heat and mass transfer with the cold reflux streams moving down the tower via the side streams. The distillate is withdrawn as LVGO and tows other cuts, MVGO, HVGO and vacuum residue. The tow cuts of MVGO and HVGO are necessary to extract heat from the tower at more advantageous level from the HVGO pumparound. Figure represents a typical process flow diagram of VDU operated under wet conditions with three vacuum gas oil (VGO) side products from light (L) to medium-boiling (M) to heavy (H) – LVGO, MVGO, and HVGO. The furnace outlet temperature varies from 380 to 420 °C, depending on the feedstock type. The products are taken off at the appropriate sections are cooled either by heat exchange with colder streams in the atmospheric unit, by air coolers or, in some cases as heating mediums to light end reboilers. They are then pumped to storage. *(A.-Fu Chang, (2012).*

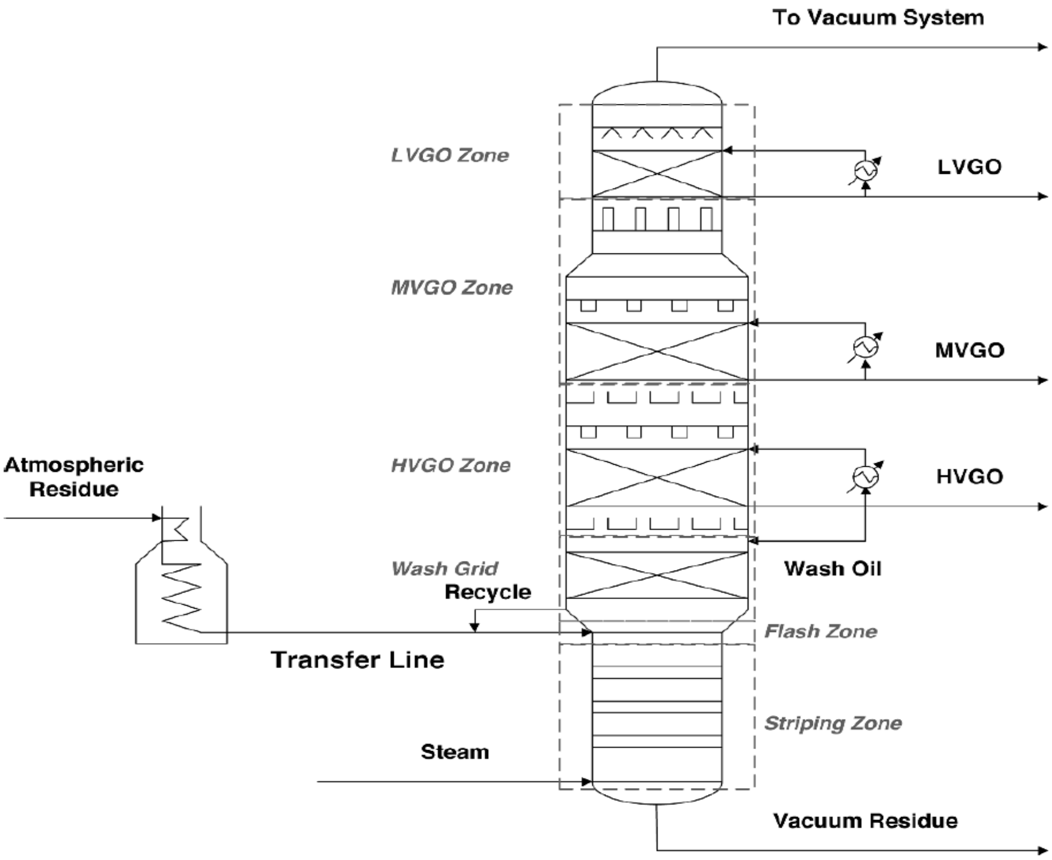


Figure 2.1: Typical Process Flow Diagram Of VDU*(A.- Fu Chang, (2012)*

2.8 Deep Cut:

The term deepcut vacuum column operation has no Standard definition, and one refiner's deepcut operation is another's typical operation. The authors' definition of deepcut is a heavy vacuum gas oil (HVGO) TBP cut point of 1100+ ºF degrees on a light crude and 1050+ ºF on high –metals crude oil. More than one definition Of HVGO cut point is used in the industry. This definition is inadequate for high cut points, and it predicts different cut points for two different types of crude vacuum column operations yielding 9 an equivalent vacuum residue yield. Here, the HVGO TBP cut point is defined as the temperature on the whole crude TBP curve Figure 2.2: that corresponds to the cumulative distillate yield. *(S.W.Golden, (Aprill 1994).*

Deep cut vacuum unit modifications improve gas oil yield. This means a higher residue initial boiling point and greater heavy vacuum gas oil (HVGO) production. The higher vacuum gas oil yield associated with deep cut improves refinery margins. The amount of additional gas oil is a function of the processing conditions and the quantity of recoverable gas oil in the feed. Cutting deeper into the bottoms to recover desirable product is not a new strategy for any fractionation system. Increased product recovery has always been beneficial to profit enhancement. *(Donald F.Scheider, (March 1997)*

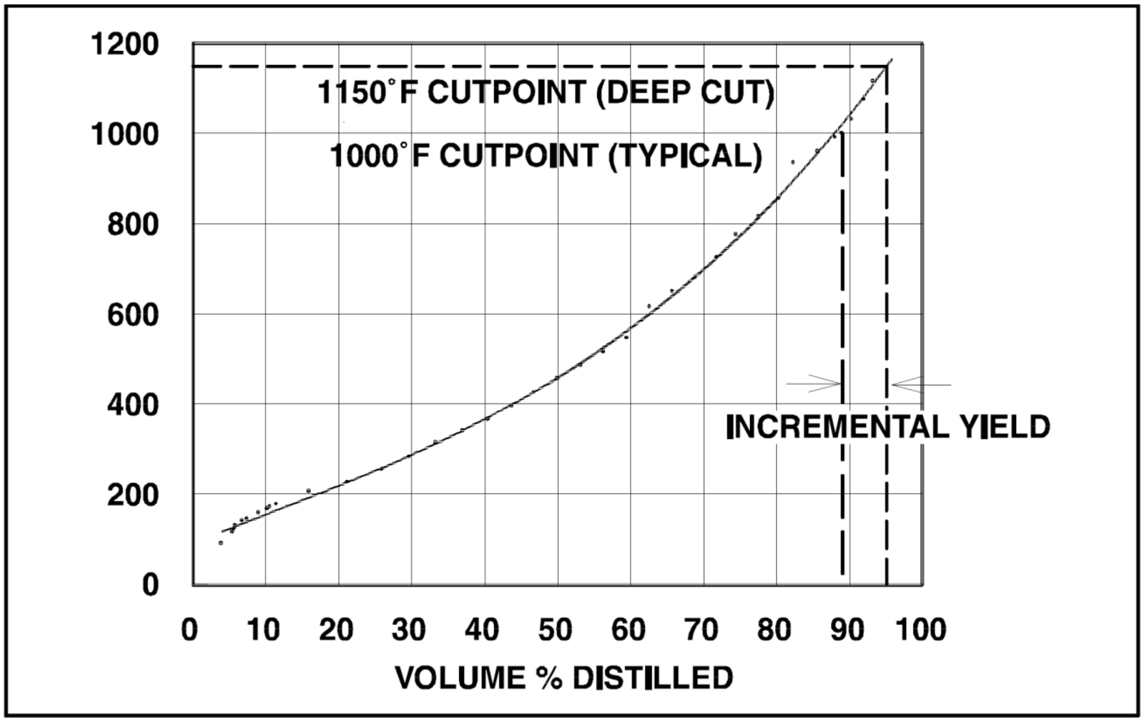


Figure 2.2: Crude Oil TBP- HVGO Cut Point *(S.W.Golden,(Aprill 1994)*

Chapter 3

Methodology

3.1 Project work:

- Selection of case study
- Process description
- Aspen HYSYS simulation
- Material & Energy balance
- Process design
- Cost estimation
- Results

3.2 Methodology:

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

3.3 Selection of case study:

Gasoline produced from Khartoum refinery is chosen as the subject for this project, the amount of gasoline production by Khartoum refinery is 3400 ton/day.

3.4 Process description:

A simplified process flow diagram for the VDU plant shown in figure3.17.The atmospheric residue is the main source of this process. The atmospheric residue is mix with steam at the mixture to increase velocity and minimize coke formation within the heater, afterward it is heated in a furnace to match the vacuum column conditions, then feed to vacuum column.

3.5 Material & Energy balance:

- Over all material balance.
- Material balance around vacuum distillation column.
- Energy balance around heater.
- Energy balance around HVGO pumparound.
- Energy balance around LVGO pumparound.

3.6 Process Design:

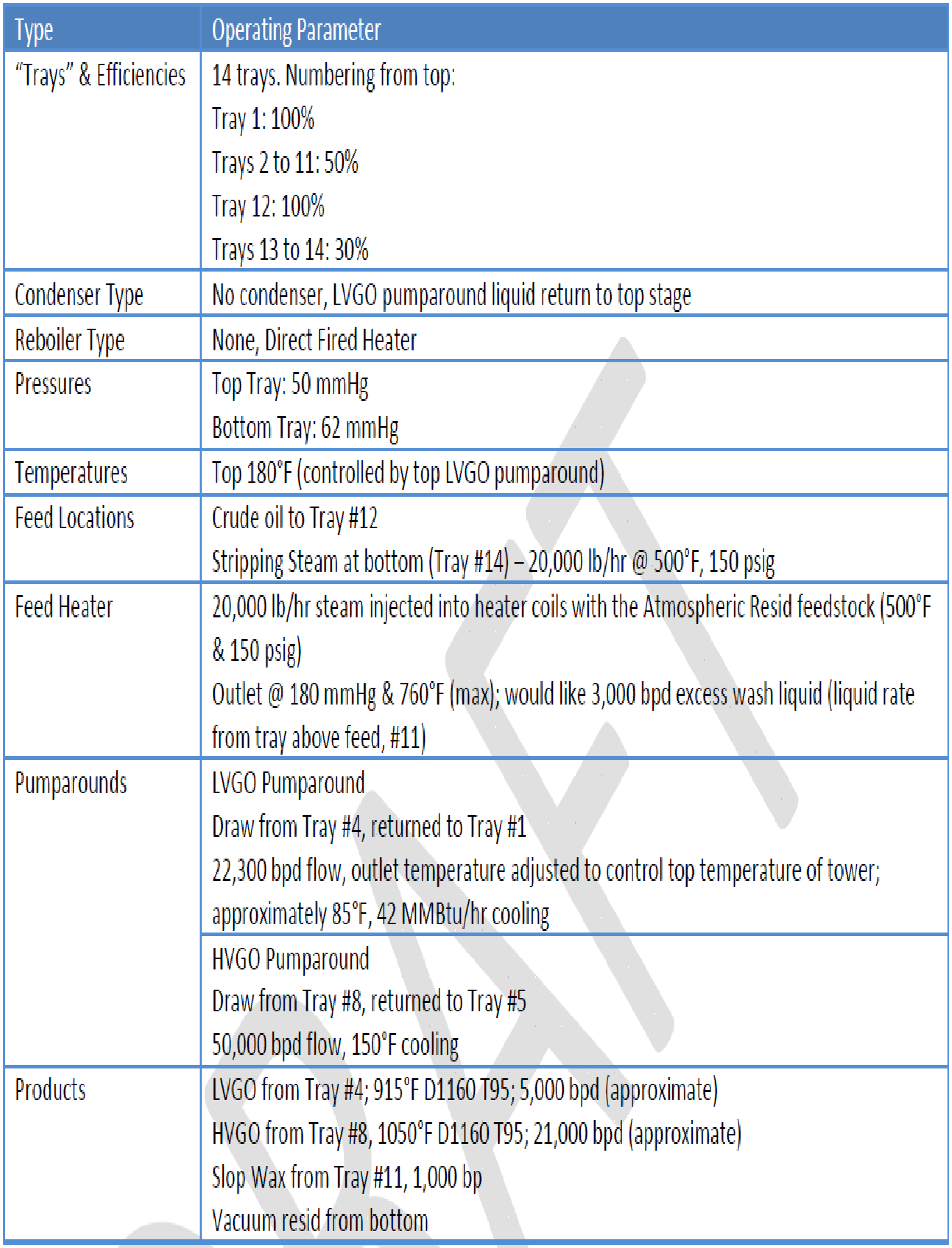
1. Determination of type of tray required.
2. Specification of the light and heavy key component.
3. Calculation of Minimum Reflux Ratio (R_{min}).
4. Calculation of Actual Reflux Ratio (R).
5. Calculation of Minimum number of stages (N_{min})
6. Calculation of theoretical number of stages (N).
7. Calculation of the column efficiency (E_o)

8. Calculation of actual number of stages (N_a).
9. Calculation of the height of the column (H_t).
10. Calculation of the feed plate location.
11. Calculation of diameter of the column (D_c).
12. Determination of entrainment correlation (Ψ).
13. Calculation of the Pressure drop.
14. Calculation of the Down comer liquid back up.
15. Calculation of the Down comer residence time.
16. Checking of flooding.

3.7 Cost estimation:

- Estimation of total capital investment.
- Determination of Fixed capital investment.
- Determination of Working Capital.
- Estimation of Total Product cost.
- Determination of Manufacturing Cost.
- Determination of Fixed Charges.
- Determination of Direct Production Cost.
- Determination of Plant overhead Costs.
- Determination of General Expenses.
- Determination of Administrative costs.
- Determination of Distribution and Selling costs.
- Determination of Research and Development costs

Table 3.1, Definitions for Vacuum Distillation Column

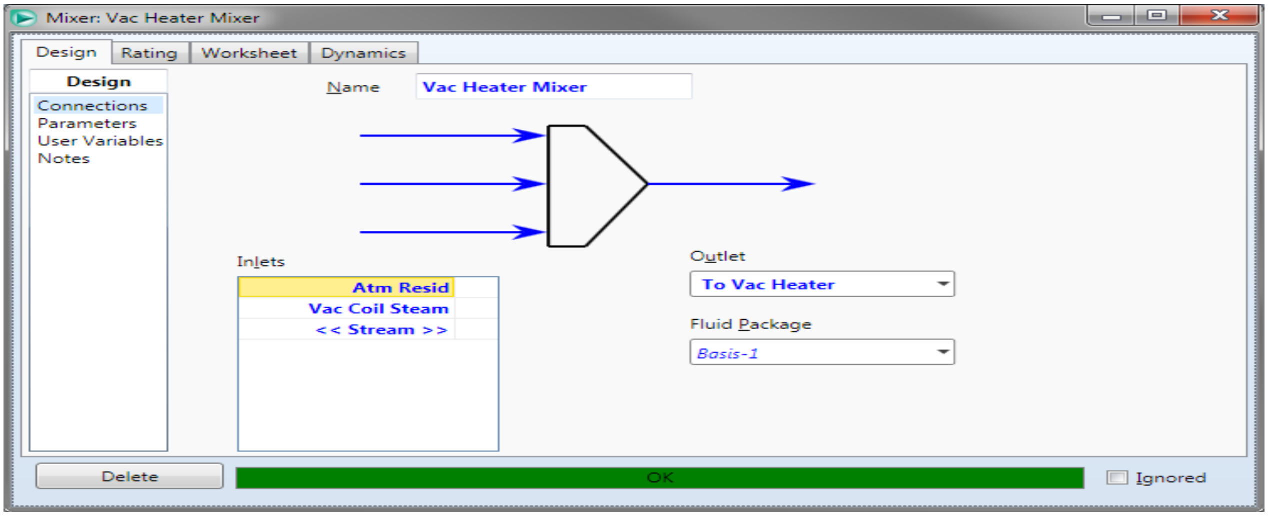


Type	Operating Parameter
"Trays" & Efficiencies	14 trays. Numbering from top: Tray 1: 100% Trays 2 to 11: 50% Tray 12: 100% Trays 13 to 14: 30%
Condenser Type	No condenser, LVGO pumparound liquid return to top stage
Reboiler Type	None, Direct Fired Heater
Pressures	Top Tray: 50 mmHg Bottom Tray: 62 mmHg
Temperatures	Top 180°F (controlled by top LVGO pumparound)
Feed Locations	Crude oil to Tray #12 Stripping Steam at bottom (Tray #14) – 20,000 lb/hr @ 500°F, 150 psig
Feed Heater	20,000 lb/hr steam injected into heater coils with the Atmospheric Resid feedstock (500°F & 150 psig) Outlet @ 180 mmHg & 760°F (max); would like 3,000 bpd excess wash liquid (liquid rate from tray above feed, #11)
Pumparounds	LVGO Pumparound Draw from Tray #4, returned to Tray #1 22,300 bpd flow, outlet temperature adjusted to control top temperature of tower; approximately 85°F, 42 MMBtu/hr cooling
	HVGO Pumparound Draw from Tray #8, returned to Tray #5 50,000 bpd flow, 150°F cooling
Products	LVGO from Tray #4; 915°F D1160 T95; 5,000 bpd (approximate) HVGO from Tray #8, 1050°F D1160 T95; 21,000 bpd (approximate) Slop Wax from Tray #11, 1,000 bp Vacuum resid from bottom

3.8 ASPEN Hysys:

3.8.1. ASPEN Hysys work procedure:

-The first step is to mix the atmospheric residue (Atm Resid) from the Atmospheric Distillation Column with the steam upstream of the Vacuum Heater. Place a Mixer on the flow sheet & define specification from table 3.1. You will have to define the steam stream from table 3.1 this shown in figure 3.1.and figure 3.2.



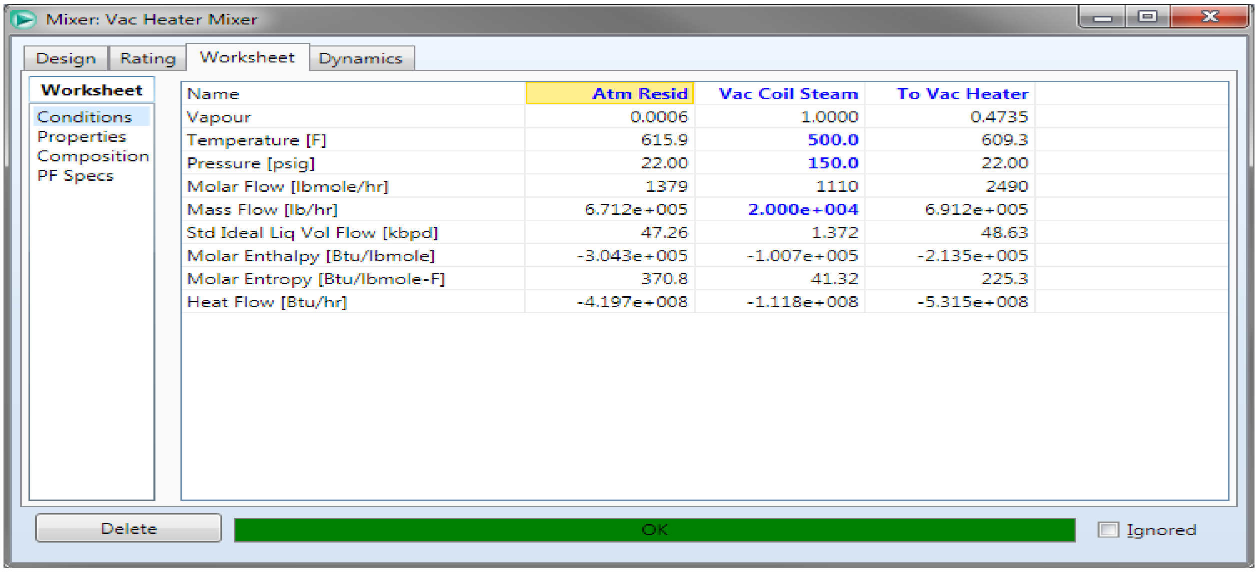
Name: Vac Heater Mixer

Inlets:
Atm Resid
Vac Coil Steam
<< Stream >>

Outlet: To Vac Heater

Fluid Package: Basis-1

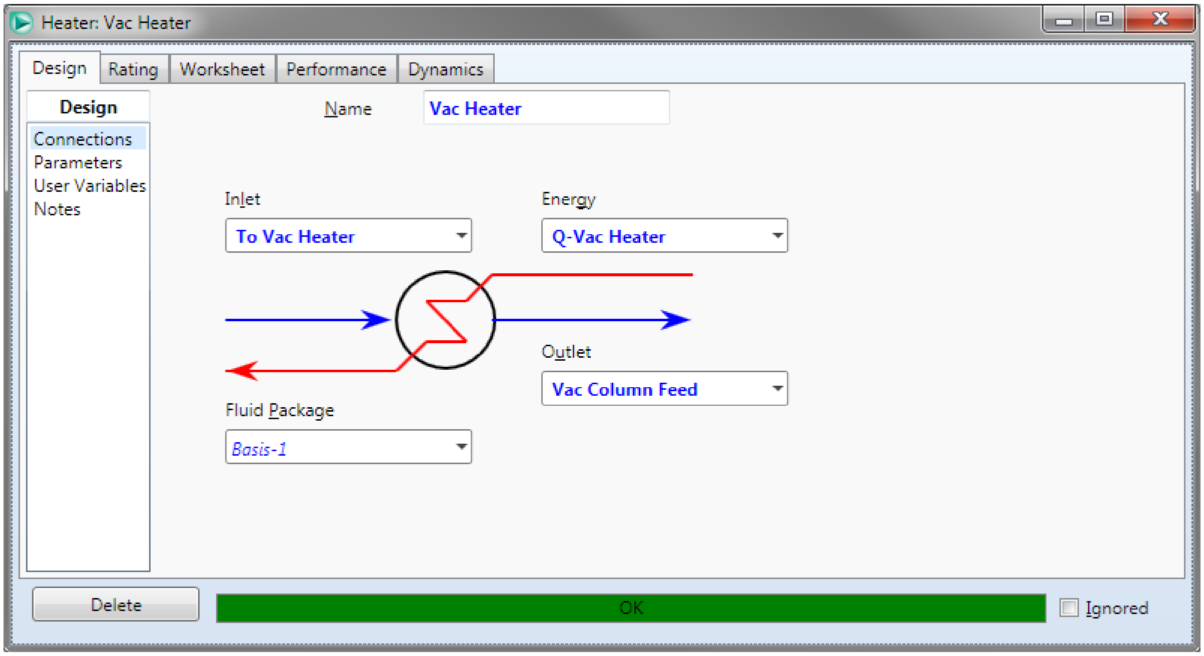
Figure 3.1: Installation of mixer



Name	Atm Resid	Vac Coil Steam	To Vac Heater
Vapour	0.0006	1.0000	0.4735
Temperature [F]	615.9	500.0	609.3
Pressure [psig]	22.00	150.0	22.00
Molar Flow [lbmole/hr]	1379	1110	2490
Mass Flow [lb/hr]	6.712e+005	2.000e+004	6.912e+005
Std Ideal Liq Vol Flow [kbpd]	47.26	1.372	48.63
Molar Enthalpy [Btu/lbmole]	-3.043e+005	-1.007e+005	-2.135e+005
Molar Entropy [Btu/lbmole-F]	370.8	41.32	225.3
Heat Flow [Btu/hr]	-4.197e+008	-1.118e+008	-5.315e+008

Figure 3.2: definition of steam stream

- Create a new Heater on the flow sheet & call it vacuum heater(Vac Heater) and define specification from table 3.1 and define vacuum column feed (vac column feed) from table 3.1 this shown in figure 3.3 and figure 3.4.



Name: Vac Heater

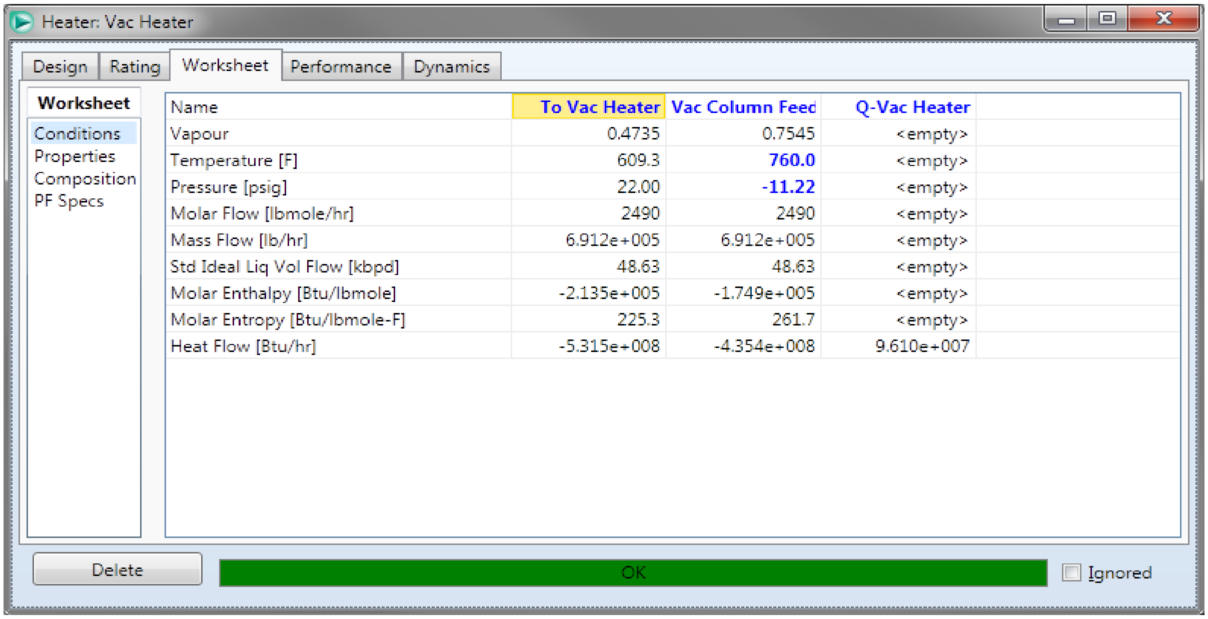
Inlet: To Vac Heater

Energy: Q-Vac Heater

Outlet: Vac Column Feed

Fluid Package: *Basis-1*

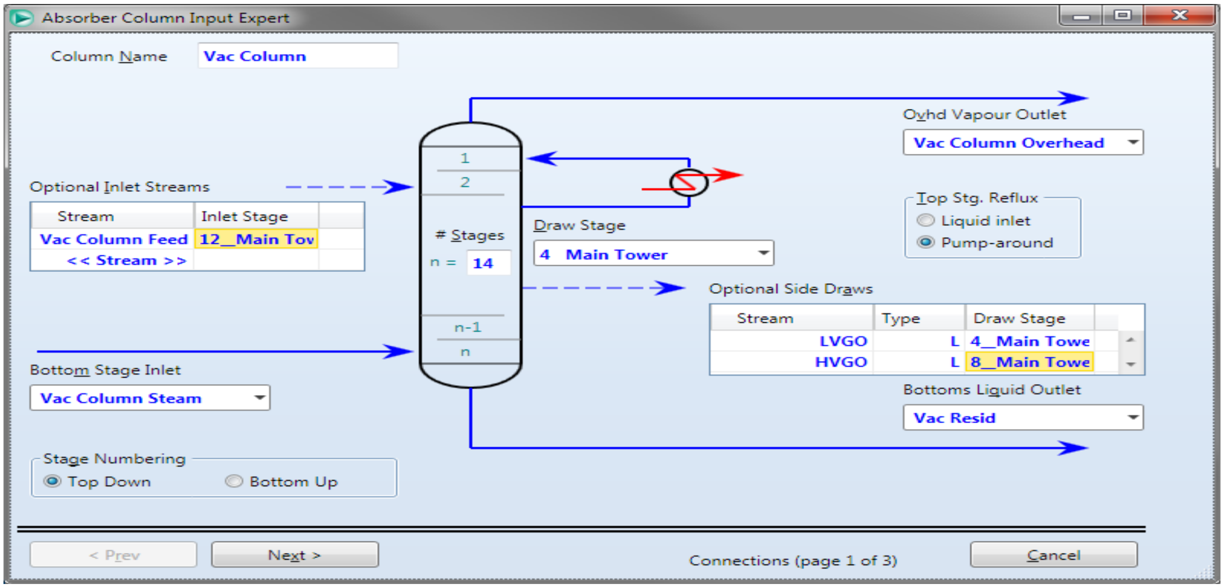
Figure 3.3: installation of Vac heater.



Name	To Vac Heater	Vac Column Feed	Q-Vac Heater
Vapour	0.4735	0.7545	<empty>
Temperature [F]	609.3	760.0	<empty>
Pressure [psig]	22.00	-11.22	<empty>
Molar Flow [lbmole/hr]	2490	2490	<empty>
Mass Flow [lb/hr]	6.912e+005	6.912e+005	<empty>
Std Ideal Liq Vol Flow [kbpd]	48.63	48.63	<empty>
Molar Enthalpy [Btu/lbmole]	-2.135e+005	-1.749e+005	<empty>
Molar Entropy [Btu/lbmole-F]	225.3	261.7	<empty>
Heat Flow [Btu/hr]	-5.315e+008	-4.354e+008	9.610e+007

Figure 3.4: definition of vac column feed.

- Create an Absorber Column on the flow sheet then double-click start to fill in the information. Fill in the basic information from table 3.1 and Specify that the top stage reflux comes from a Pump-around. Specify the LVGO, HVGO, & Slop Wax streams on this form as Optional Side Draws this shown in figure 3.5.



Column Name: Vac Column

Optional Inlet Streams

Stream	Inlet Stage
Vac Column Feed	12__Main Tov
<< Stream >>	

Stages n = 14

Draw Stage: 4 Main Tower

Ovhd Vapour Outlet: Vac Column Overhead

Top Stg. Reflux
○ Liquid inlet
● Pump-around

Optional Side Draws

Stream	Type	Draw Stage
LVGO	L	4__Main Towe
HVGO	L	8__Main Towe

Bottom Stage Inlet: Vac Column Steam

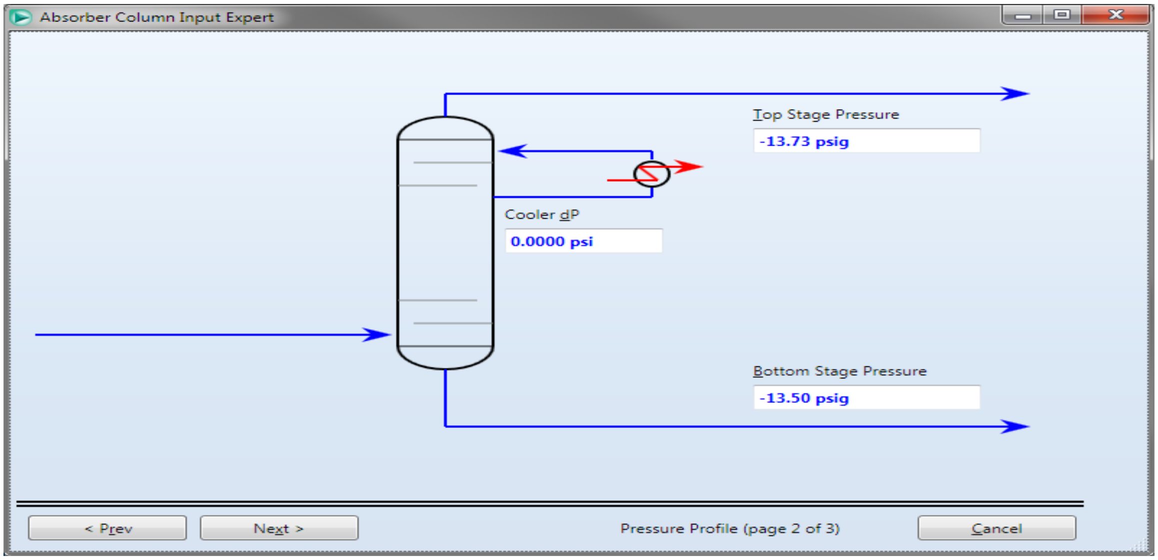
Bottoms Liquid Outlet: Vac Resid

Stage Numbering
● Top Down ○ Bottom Up

Connections (page 1 of 3)

Figure 3.5: Installation of Absorber Column.

- Initialize the pressure profile the specification found in table 3.1 this shown in figure 3.6.



Top Stage Pressure: -13.73 psig

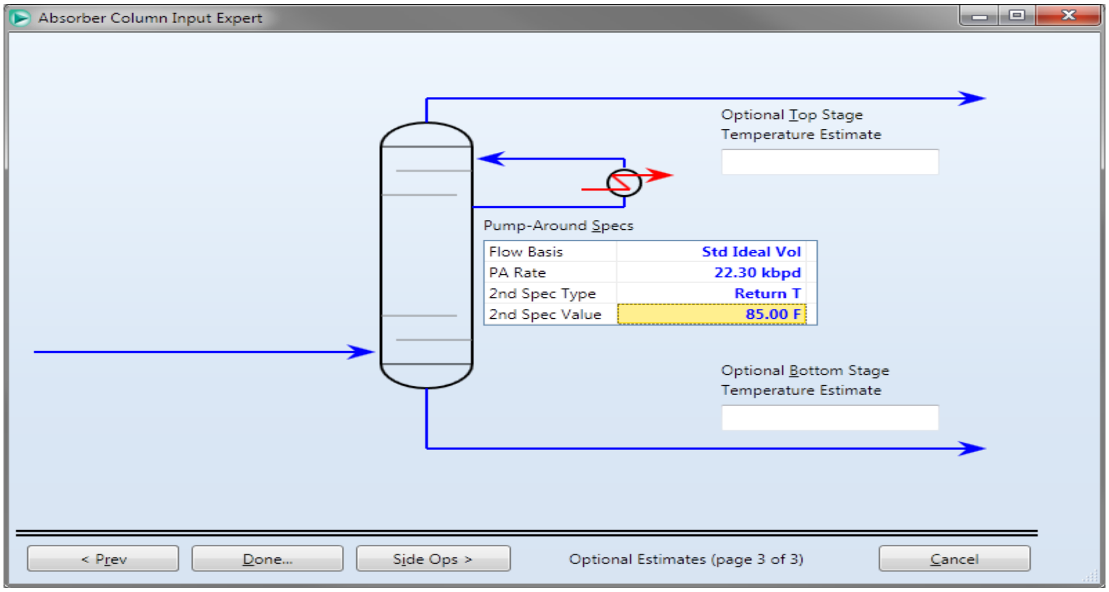
Cooler dP: 0.0000 psi

Bottom Stage Pressure: -13.50 psig

Pressure Profile (page 2 of 3)

Figure 3.6: Definition of pressure.

- Add temperature estimates & flow information for the top pumparound. Enter the data for the light vacuum gas oil(LVGO)Pumparound from table 3.1.

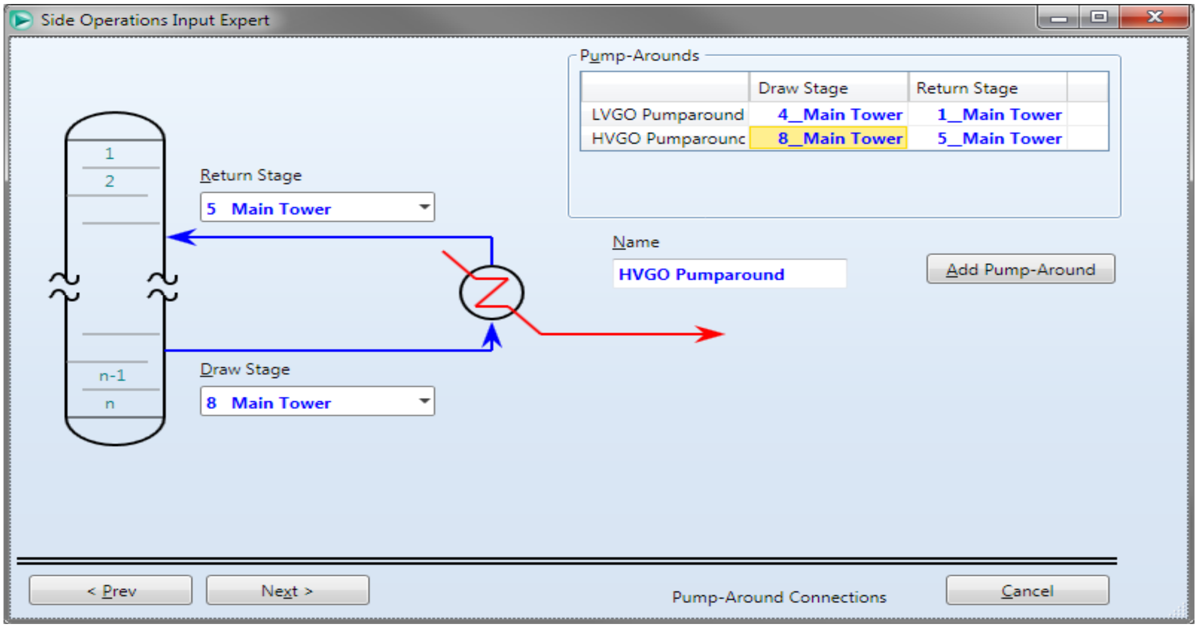


Pump-Around Specs

Flow Basis	Std Ideal Vol
PA Rate	22.30 kbpd
2nd Spec Type	Return T
2nd Spec Value	85.00 F

Figure 3.7: Definition of LVGO pumparound.

- Changed the Name from the default to light vacuum gas oil (LVGO) Pumparound. Then click Add Pump-Around and define the heavy vacuum gas oil (HVGO) Pumparound.



Pump-Arounds

	Draw Stage	Return Stage
LVGO Pumparound	4__Main Tower	1__Main Tower
HVGO Pumparounc	8__Main Tower	5__Main Tower

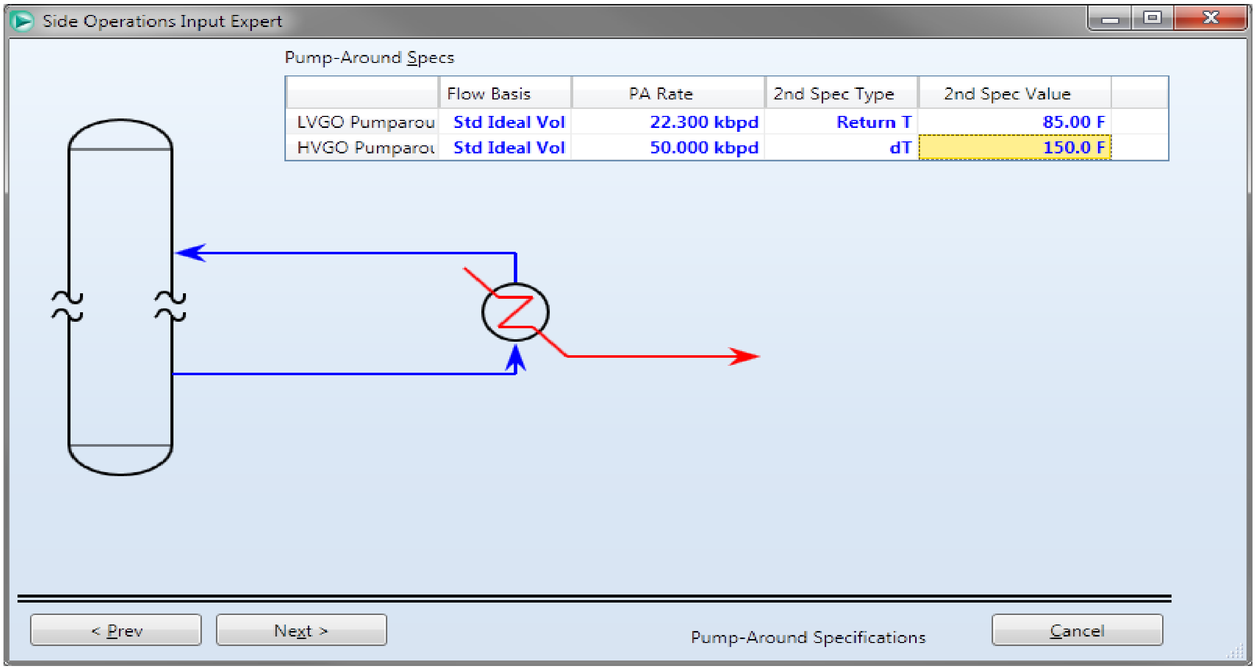
Return Stage: 5 Main Tower

Draw Stage: 8 Main Tower

Name: HVGO Pumparound

Figure 3.8: Definition of LVGO pumparound and HVGO pumparound.

- Add the heavy vacuum gas oil (HVGO) Pumparound specs from table 3.1 this shown in figure 3.9.



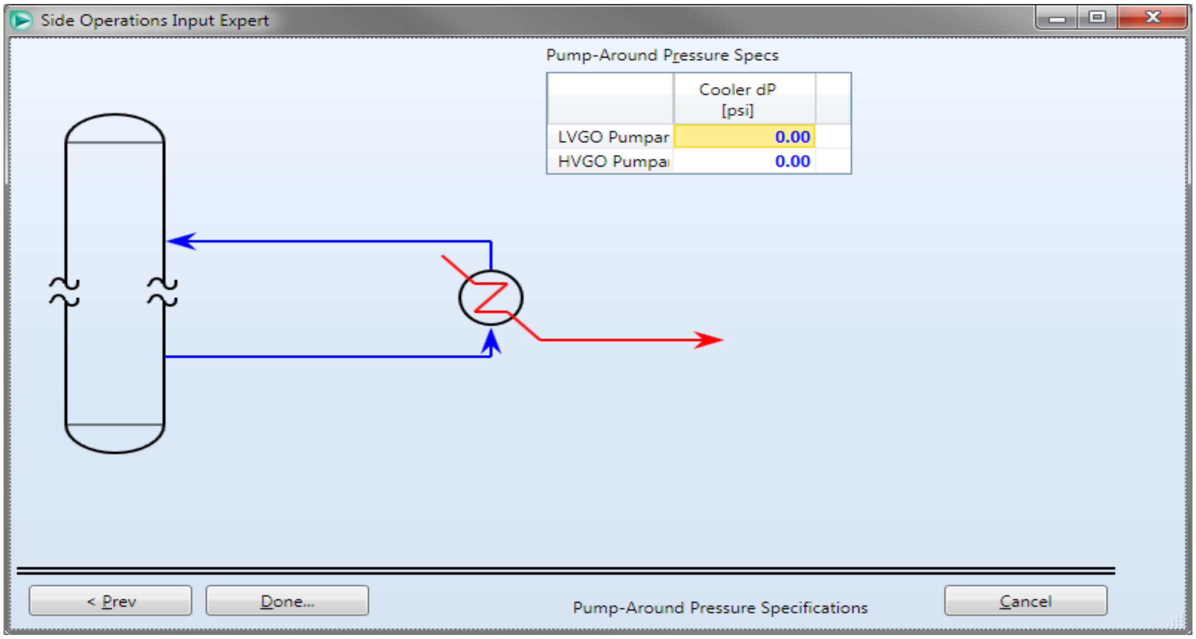
Pump-Around Specs

	Flow Basis	PA Rate	2nd Spec Type	2nd Spec Value
LVGO Pumparou	Std Ideal Vol	22.300 kbpd	Return T	85.00 F
HVGO Pumparou	Std Ideal Vol	50.000 kbpd	dT	150.0 F

Pump-Around Specifications

Figure 3.9: Definition of HVGO pumparound.

- Define pressure drops through the pumparounds this shown in figure 3.10.



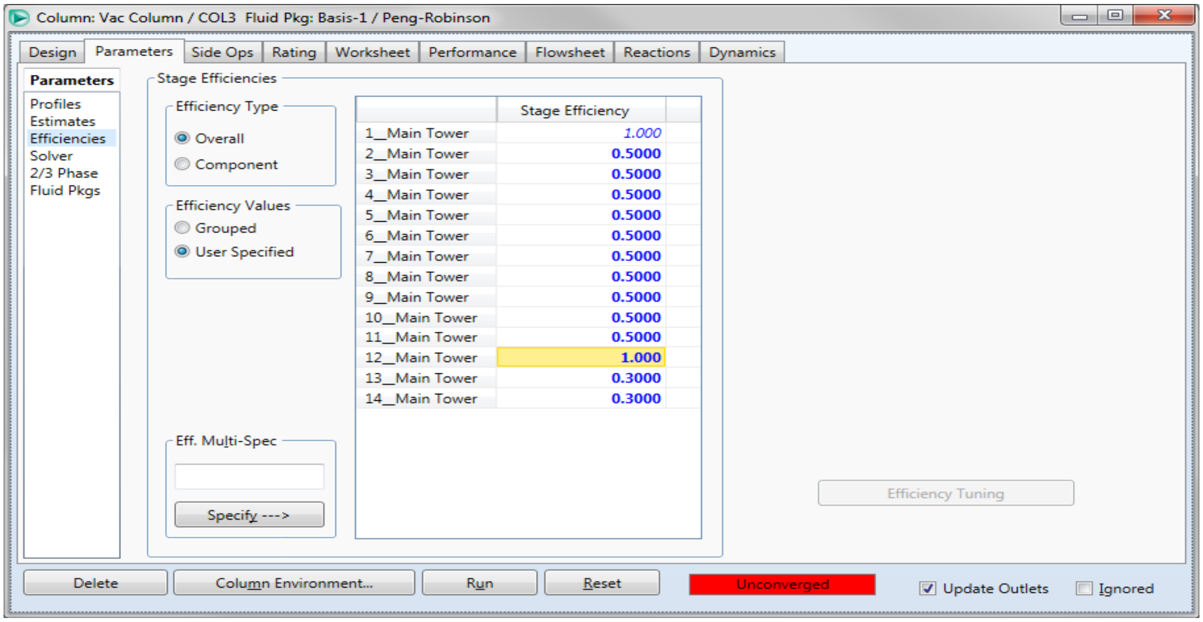
Pump-Around Pressure Specs

	Cooler dP [psi]
LVGO Pumpar	0.00
HVGO Pumpa	0.00

Pump-Around Pressure Specifications

Figure 3.10: Definition of pressure drops.

- Enter the efficiencies for the stages. Select Efficiencies under the Parameters tab and enter the values from Table 3.1 this shown in figure 3.11.



Efficiency Type: ● Overall ○ Component

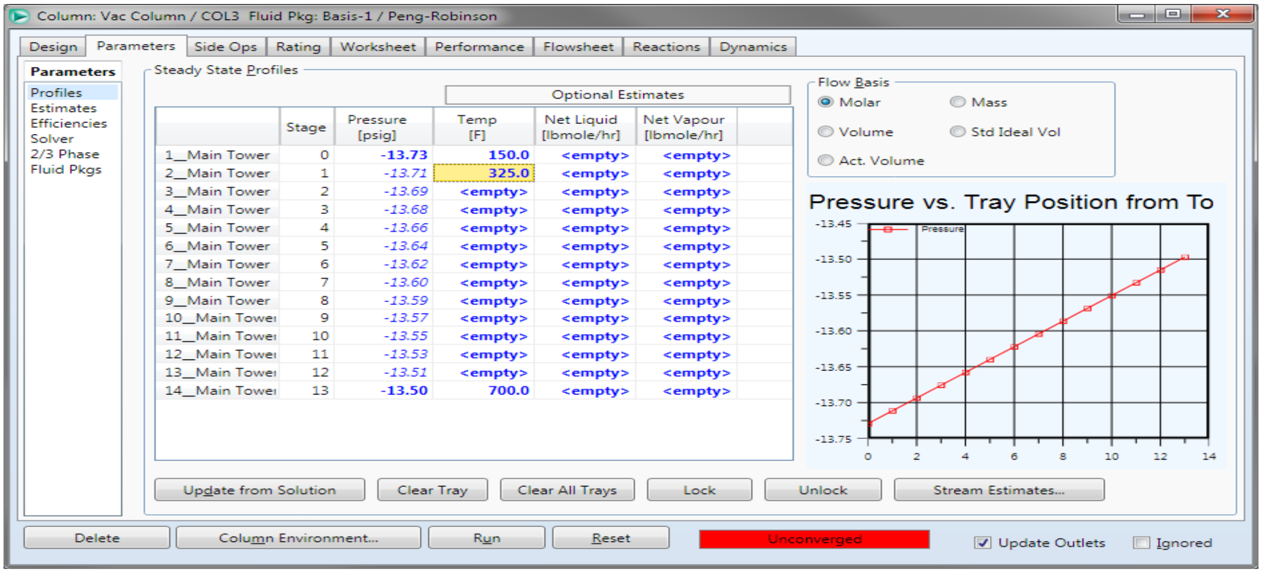
Efficiency Values: ○ Grouped ● User Specified

	Stage Efficiency
1__Main Tower	1.000
2__Main Tower	0.5000
3__Main Tower	0.5000
4__Main Tower	0.5000
5__Main Tower	0.5000
6__Main Tower	0.5000
7__Main Tower	0.5000
8__Main Tower	0.5000
9__Main Tower	0.5000
10__Main Tower	0.5000
11__Main Tower	0.5000
12__Main Tower	1.000
13__Main Tower	0.3000
14__Main Tower	0.3000

☑ Update Outlets ☐ Ignored

Figure 3.11: Input efficiencies values.

- Go to the Profiles item under the Parameters tab. It’s pretty typical to have a top temperature of about 150°F and a bottom temperature of 700°F. You may also want to specify the 2nd stage temperature of 325°F this shown in figure 3.12.



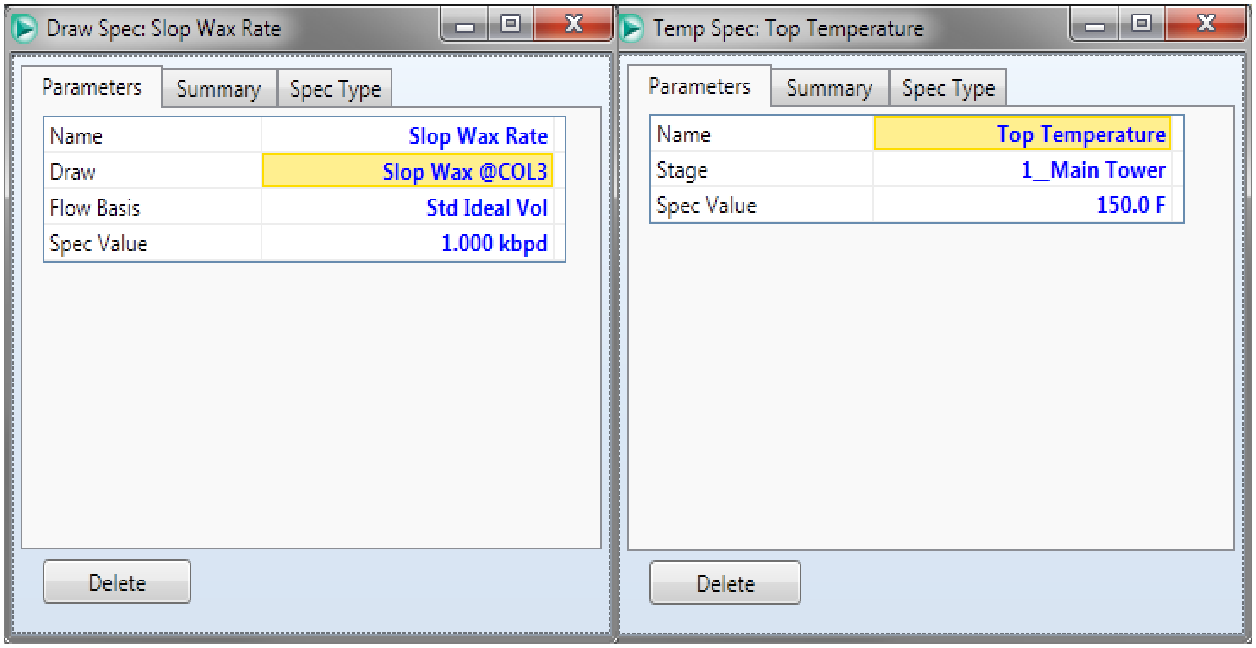
Flow Basis: ● Molar ○ Mass ○ Volume ○ Std Ideal Vol ○ Act. Volume

	Stage	Pressure [psig]	Temp [F]	Net Liquid [lbmole/hr]	Net Vapour [lbmole/hr]
			Optional Estimates		
1__Main Tower	0	-13.73	150.0	<empty>	<empty>
2__Main Tower	1	-13.71	325.0	<empty>	<empty>
3__Main Tower	2	-13.69	<empty>	<empty>	<empty>
4__Main Tower	3	-13.68	<empty>	<empty>	<empty>
5__Main Tower	4	-13.66	<empty>	<empty>	<empty>
6__Main Tower	5	-13.64	<empty>	<empty>	<empty>
7__Main Tower	6	-13.62	<empty>	<empty>	<empty>
8__Main Tower	7	-13.60	<empty>	<empty>	<empty>
9__Main Tower	8	-13.59	<empty>	<empty>	<empty>
10__Main Towei	9	-13.57	<empty>	<empty>	<empty>
11__Main Towei	10	-13.55	<empty>	<empty>	<empty>
12__Main Towei	11	-13.53	<empty>	<empty>	<empty>
13__Main Towei	12	-13.51	<empty>	<empty>	<empty>
14__Main Towei	13	-13.50	700.0	<empty>	<empty>

☑ Update Outlets ☐ Ignored

Figure 3.12: Input temperature value.

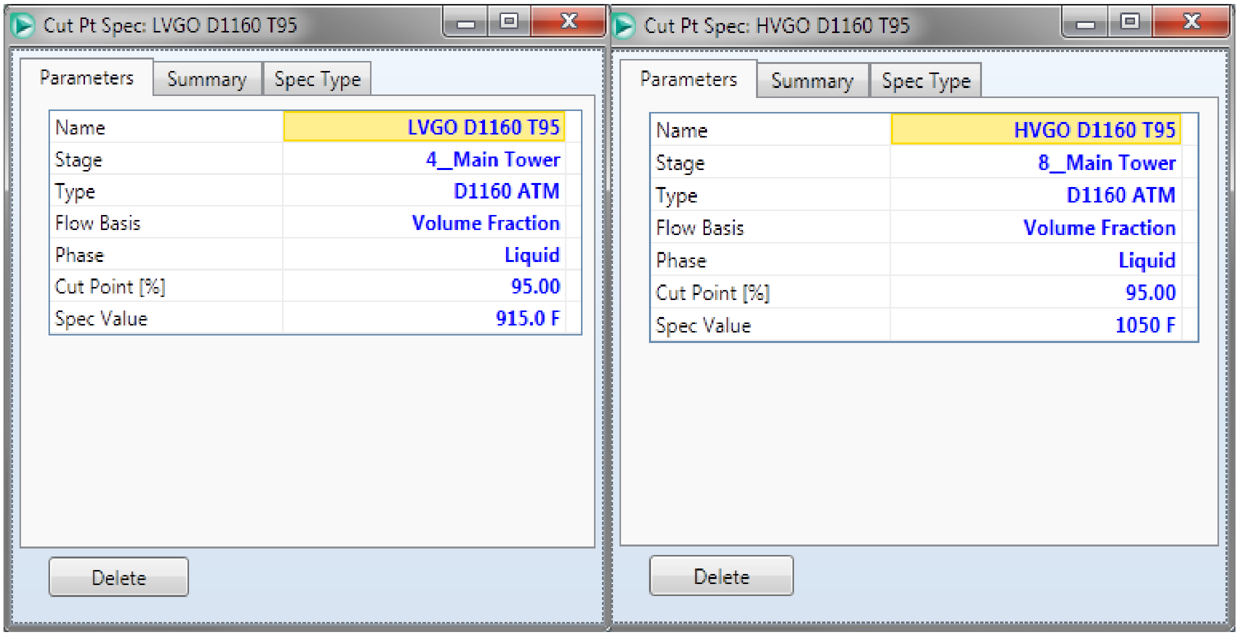
- Select the Specs item in the left-hand column under the Design tab and add spec Slop Wax Rate, top temperature spec, D1160 specs for the LVGO and HVGO, Net from #11 and Net from #4 all of this specification found in table 3.1 shown in figure 3.13, figure 3.14, figure 3.15 and overall specs shown in figure 3.16.



Name	Slop Wax Rate
Draw	Slop Wax @COL3
Flow Basis	Std Ideal Vol
Spec Value	1.000 kbpd

Name	Top Temperature
Stage	1__Main Tower
Spec Value	150.0 F

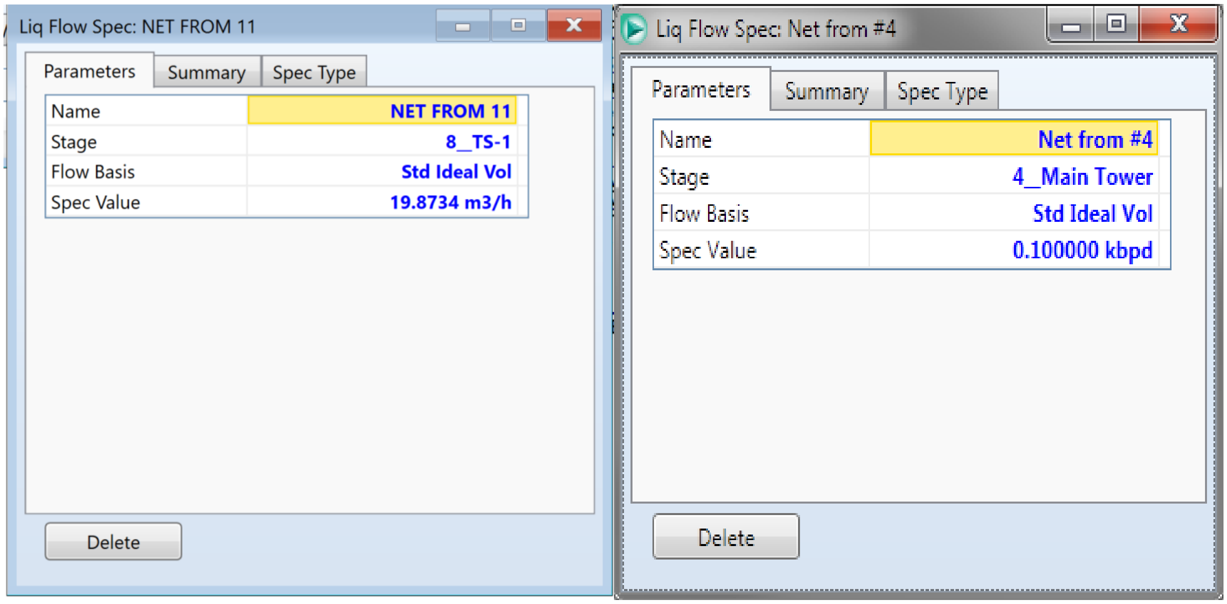
Figure 3.13: Definition of Slop Wax Rate and top temperature spec.



Name	LVGO D1160 T95
Stage	4__Main Tower
Type	D1160 ATM
Flow Basis	Volume Fraction
Phase	Liquid
Cut Point [%]	95.00
Spec Value	915.0 F

Name	HVGO D1160 T95
Stage	8__Main Tower
Type	D1160 ATM
Flow Basis	Volume Fraction
Phase	Liquid
Cut Point [%]	95.00
Spec Value	1050 F

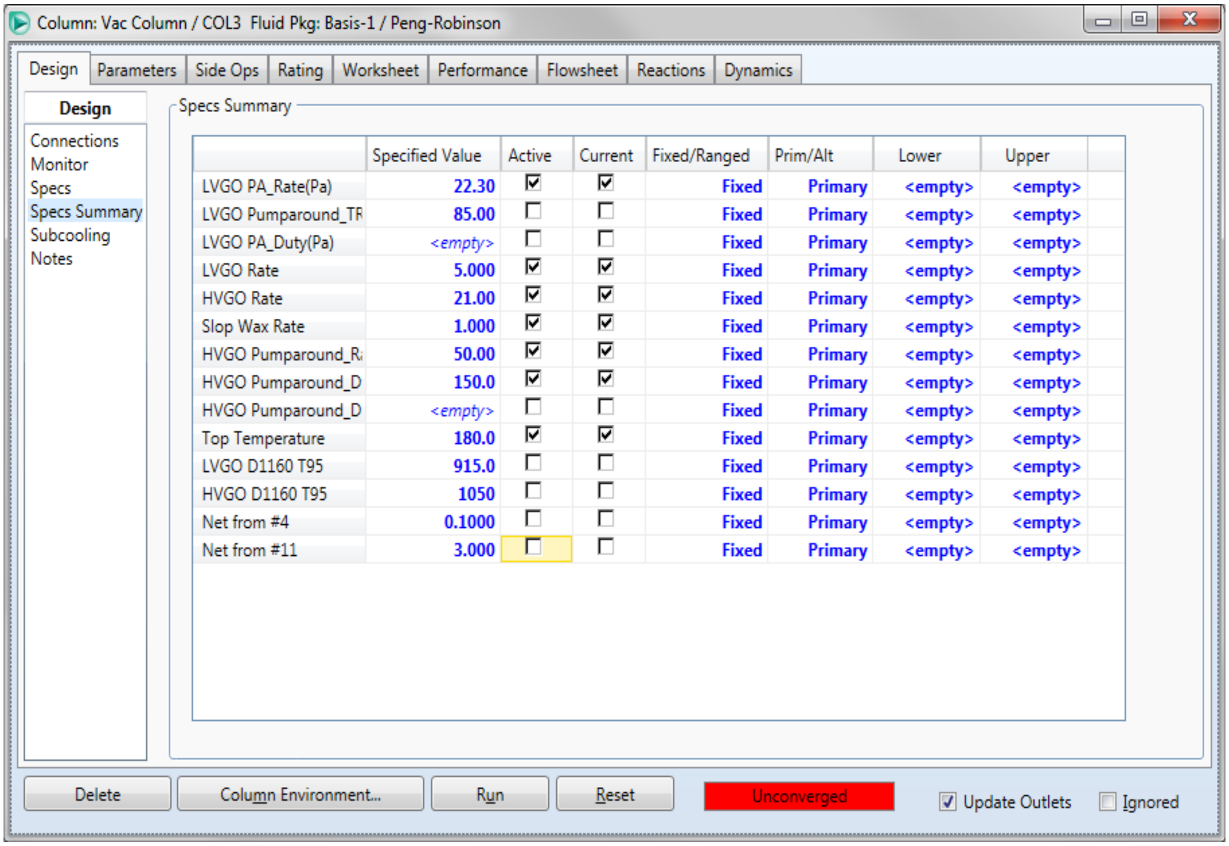
Figure 3.14: Definition of D1160 specs for the LVGO and HVGO.



Name	NET FROM 11
Stage	8__TS-1
Flow Basis	Std Ideal Vol
Spec Value	19.8734 m3/h

Name	Net from #4
Stage	4__Main Tower
Flow Basis	Std Ideal Vol
Spec Value	0.100000 kbpd

Figure 3.15: Definition of Net from #11 and Net from #4.



	Specified Value	Active	Current	Fixed/Ranged	Prim/Alt	Lower	Upper
LVGO PA_Rate(Pa)	22.30	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO Pumparound_TR	85.00	☐	☐	Fixed	Primary	<empty>	<empty>
LVGO PA_Duty(Pa)	<empty>	☐	☐	Fixed	Primary	<empty>	<empty>
LVGO Rate	5.000	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO Rate	21.00	☑	☑	Fixed	Primary	<empty>	<empty>
Slop Wax Rate	1.000	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO Pumparound_R	50.00	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO Pumparound_D	150.0	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO Pumparound_D	<empty>	☐	☐	Fixed	Primary	<empty>	<empty>
Top Temperature	180.0	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO D1160 T95	915.0	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO D1160 T95	1050	☐	☐	Fixed	Primary	<empty>	<empty>
Net from #4	0.1000	☐	☐	Fixed	Primary	<empty>	<empty>
Net from #11	3.000	☐	☐	Fixed	Primary	<empty>	<empty>

Figure 3.16: Overall specs.

-The final model of vacuum distillation unit shown in figure 3.17.

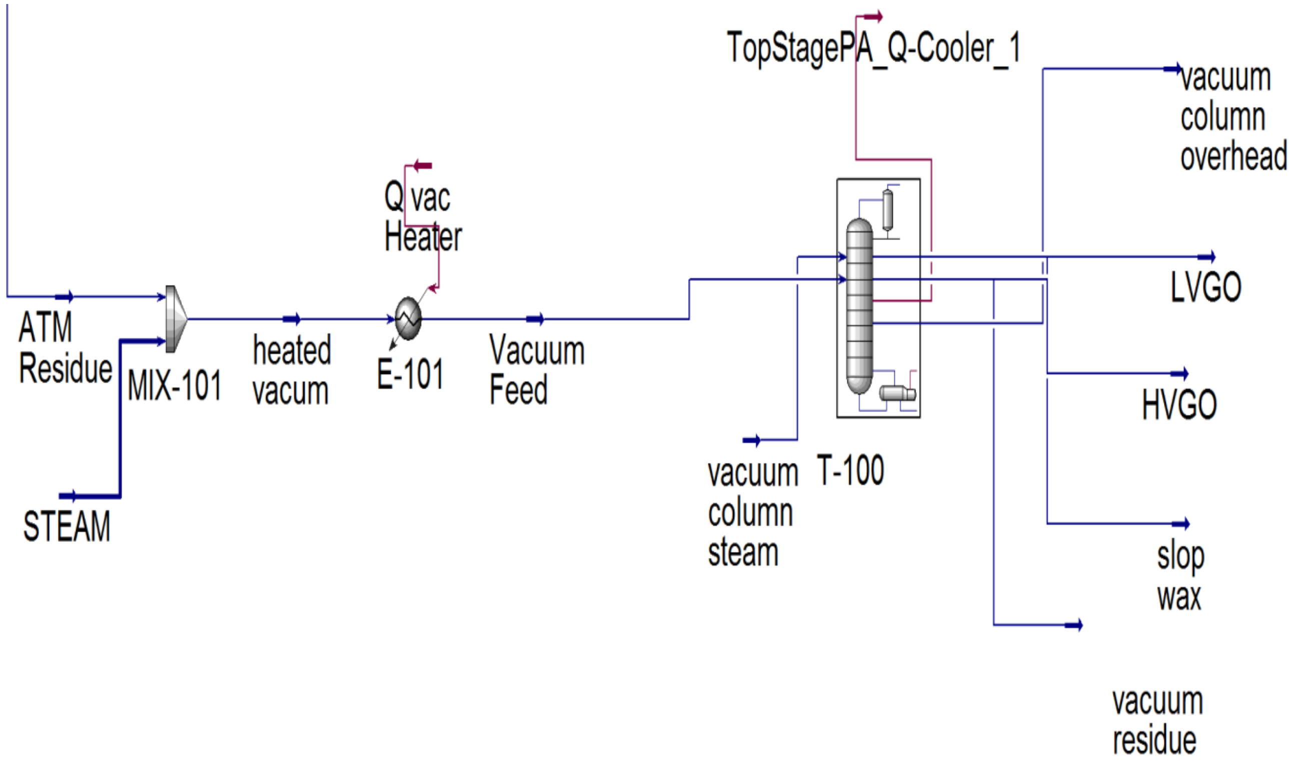
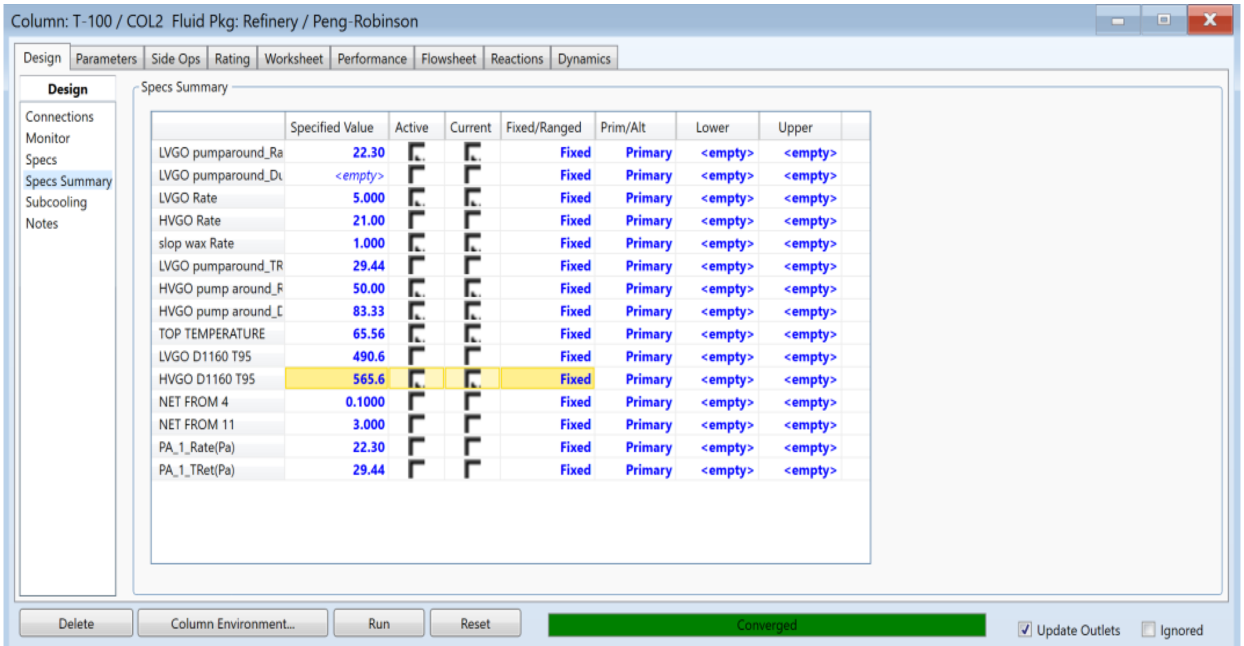


Figure 3.17: Overall view of vacuum distillation process.

3.8.2. Improving production of product:

By changing of specification on specs summary this improve the rate of product. We use HVGO D1160 T95 specs to improve the production of HVGO this shown in figure 4.18and useNET FROM #4 specs to improve the production of LVGO in figure 3.19.

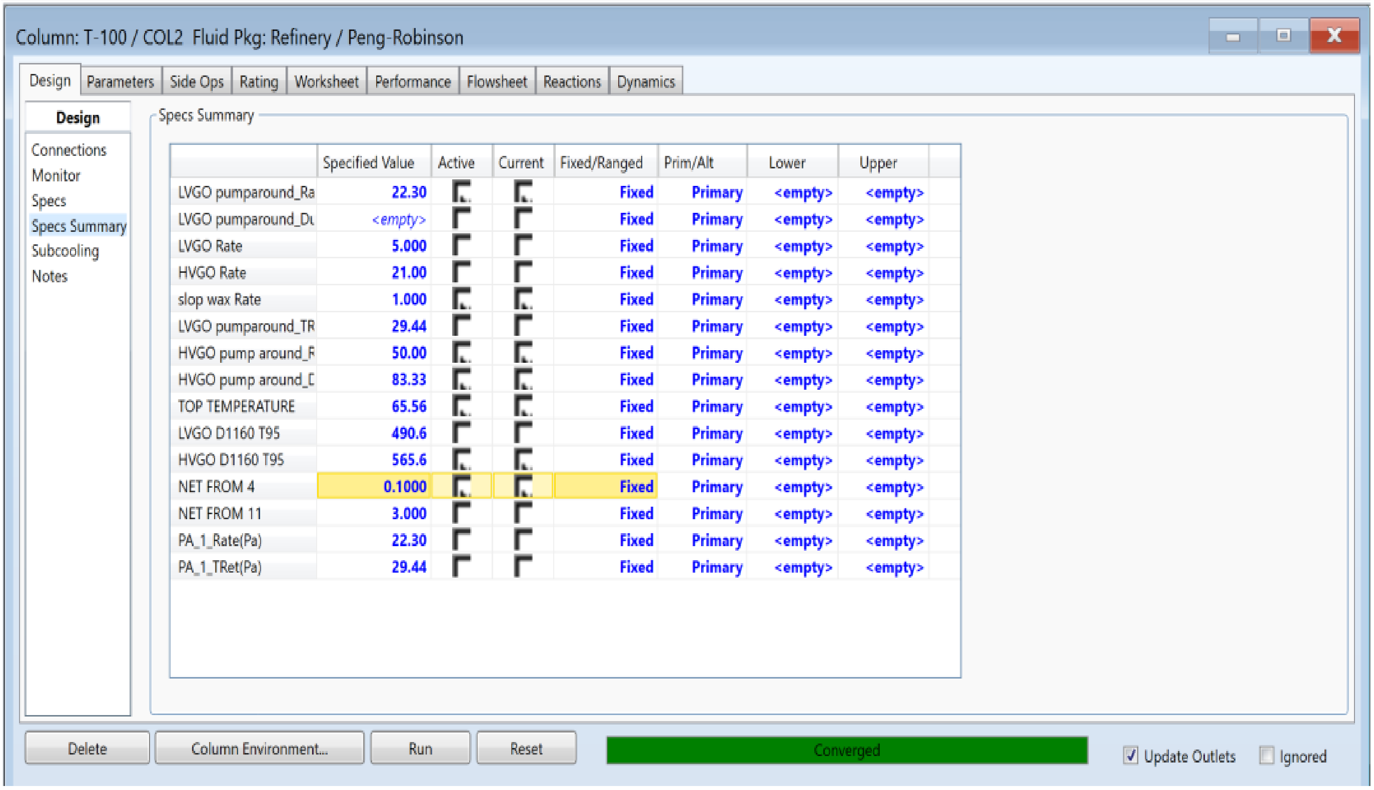


Column: T-100 / COL2 Fluid Pkg: Refinery / Peng-Robinson

Specs Summary

	Specified Value	Active	Current	Fixed/Ranged	Prim/Alt	Lower	Upper
LVGO pumparound_Ra	22.30	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO pumparound_Du	<empty>	☐	☐	Fixed	Primary	<empty>	<empty>
LVGO Rate	5.000	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO Rate	21.00	☐	☐	Fixed	Primary	<empty>	<empty>
slop wax Rate	1.000	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO pumparound_TR	29.44	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO pump around_R	50.00	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO pump around_D	83.33	☑	☑	Fixed	Primary	<empty>	<empty>
TOP TEMPERATURE	65.56	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO D1160 T95	490.6	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO D1160 T95	565.6	☑	☑	Fixed	Primary	<empty>	<empty>
NET FROM 4	0.1000	☐	☐	Fixed	Primary	<empty>	<empty>
NET FROM 11	3.000	☐	☐	Fixed	Primary	<empty>	<empty>
PA_1_Rate(Pa)	22.30	☐	☐	Fixed	Primary	<empty>	<empty>
PA_1_TRet(Pa)	29.44	☐	☐	Fixed	Primary	<empty>	<empty>

Figure 3.18: Activation ofHVGO D1160 T95 specs



Column: T-100 / COL2 Fluid Pkg: Refinery / Peng-Robinson

Specs Summary

	Specified Value	Active	Current	Fixed/Ranged	Prim/Alt	Lower	Upper
LVGO pumparound_Ra	22.30	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO pumparound_Du	<empty>	☐	☐	Fixed	Primary	<empty>	<empty>
LVGO Rate	5.000	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO Rate	21.00	☐	☐	Fixed	Primary	<empty>	<empty>
slop wax Rate	1.000	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO pumparound_TR	29.44	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO pump around_R	50.00	☑	☑	Fixed	Primary	<empty>	<empty>
HVGO pump around_D	83.33	☑	☑	Fixed	Primary	<empty>	<empty>
TOP TEMPERATURE	65.56	☑	☑	Fixed	Primary	<empty>	<empty>
LVGO D1160 T95	490.6	☐	☐	Fixed	Primary	<empty>	<empty>
HVGO D1160 T95	565.6	☑	☑	Fixed	Primary	<empty>	<empty>
NET FROM 4	0.1000	☑	☑	Fixed	Primary	<empty>	<empty>
NET FROM 11	3.000	☐	☐	Fixed	Primary	<empty>	<empty>
PA_1_Rate(Pa)	22.30	☐	☐	Fixed	Primary	<empty>	<empty>
PA_1_TRet(Pa)	29.44	☐	☐	Fixed	Primary	<empty>	<empty>

Figure 3.19: Activation of NET FROM #4

3.9 corrosion protection:

3.9.1 Introduction:

Corrosion is a major issue in distillation equipment even with proper designs. Multiple factors can interact and create corrosive attack. With the current run length of plants between maintenance outages approaching five years, corrosion control is a must to maintain distillation efficiency and recovery.

Areas of corrosion in distillation include; crude distillation, vacuum distillation, and solvent extraction. Proper metallurgy selection and then proper chemical treatment is essential to prevent corrosion in the distillation equipment for hydrocarbon and chemicals processing.

Corrosion treatment chemicals include neutralizers, filmers, and other corrosion inhibitors. These chemical can prevent or mitigate damage from galvanic bi-metallic, aqueous acidic, and under-deposit corrosion, as well as pitting.

Corrosion rates on these internal surfaces depend on the design of the equipment, materials, atmospheric/gaseous conditions, relative humidity (RH), presence of aqueous solutions, temperature and the frequency of temperature changes, corrosion condition of the internal surfaces.

To exclude this situation were created two types of corrosion protection systems that combine dehumidification (DH) of environment applying VCI in cases of presence of vapour condensed layers in some parts of the equipment due to fluctuation the temperature. In most cases the selected inhibitor composition that in vapour space works as VCI when absorbed in condensed water layer or in aqueous solution became SCI. In this case application of the system became very simple (*Y. I. Kuznetsov, 1996*).

3.9.2 Type of corrosion protection:

a. DH/VCI
b. DVS-1
c. DVS-2

3.9.2.1 DVS-1:

Is a packaging system (Figure 3) to control external and internal corrosion in the vapour space and in the condensed on the metal surface water or aqueous solutions. This is a two layer system that allows controlling RH inside of the first layer (VCI or plain film) and excluding the entry of moisture into the inner layer from outside through the external barrier packaging film. This system recommended in most cases for corrosion protection of external surface of any size of equipment and internal surface for not very large and complicated equipment.

This system has been applied in many cases where existing packaging systems not efficient. The storage and shipment environment was extremely aggressive: RH of up to 100%, temperature from +40°C to –20°C, sea atmosphere, and industrial environment. For most of thisequipment,only DVS-1 system is efficient and cost effective. (*E. Y. Lyublinski,,(2009*)).

3.9.2.2 DVS-2 System:

Is developed for corrosion protection of different type of large and complicated design of mothballing equipment.The main advantages and disadvantages of different methods are described in Table

Table.3.2: The advantages and disadvantages of the existing methods for corrosion protection of large enclosures

Protection method	Advantages	Disadvantages
1. Replacement of existing environment with nitrogen (N2)	1. 100% efficiency in case if it is possible to achieve the fully sealing of enclosure	It is impossible to check that the environment was not fully replaced Not efficient in spaces where exist liquid that cannot be replaced with nitrogen condensed For not fully sealed enclosure the consumption of N2 is unpredictable and corrosion protection cannot be guarantied Cost of the delivery system and N2 is high
2. Applying VCI	1. Average efficiency for	1. Efficiency decreases if the concentration

	about 80% 2. Applicable in different environment 3. Low cost	decrease near than the protection concentration 2. Type of inhibitor depends on environment condition and type of metals 3. Does not exist the application technology for large and complicated equipment
3. Dehumidification	High efficient and low cost	High efficient only in fully sealed enclosures

3.9.2.3 DH/VCI System

This system consists of:

- A closed system for dehumidification and/or delivery of VCI on external and internal surfaces.
- A “pump” that periodically performs dehumidification and then delivers VCI

This is the only/unique system that allows achieving corrosion protection of complicated and very large enclosures that can contain parts from different metals. To achieve the corrosion protection of any mothballing equipment it is not necessary to know the type of metals used in the enclosures, configuration, size and condition of the metal surfaces. To achieve corrosion protection, this system includes two steps:

Step1: First step include dehumidification that allow achieving the required RH less than 50% (this is the criteria of achieving corrosion protection in vapour space). In most cases the applied RH is close to 30%).

Step2: Second step include heating of powder of liquid VCI compounds to increase their vapour pressure and delivery VCI by using the dehumidification part of the system.

The temperature range in the system allows very fast to achieve the required vapour pressure for different VCI compounds. In some cases can be used compounds that only by heating became VCI. This is the additional advantage of this system.

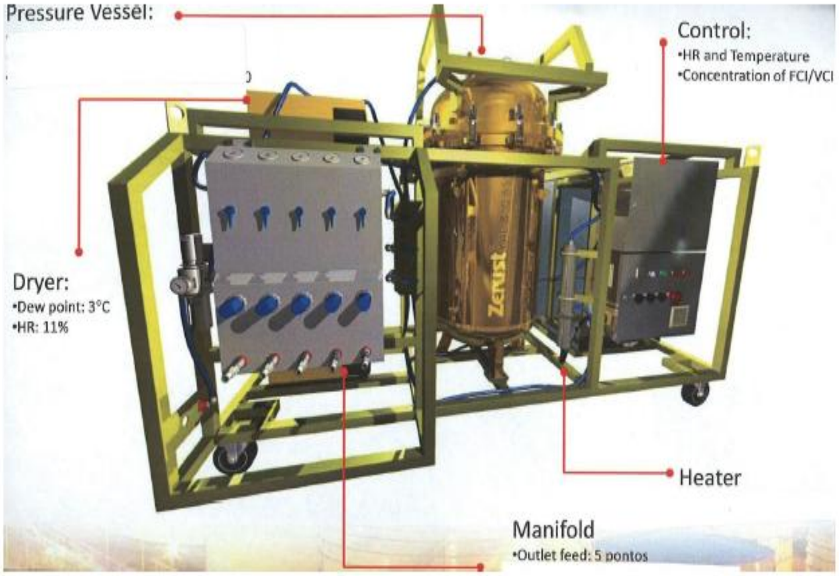


Figure 3.20: DVS-2 systems for corrosion protection of large and complicated enclosures of equipment, (E. Lyublinski, (2015))

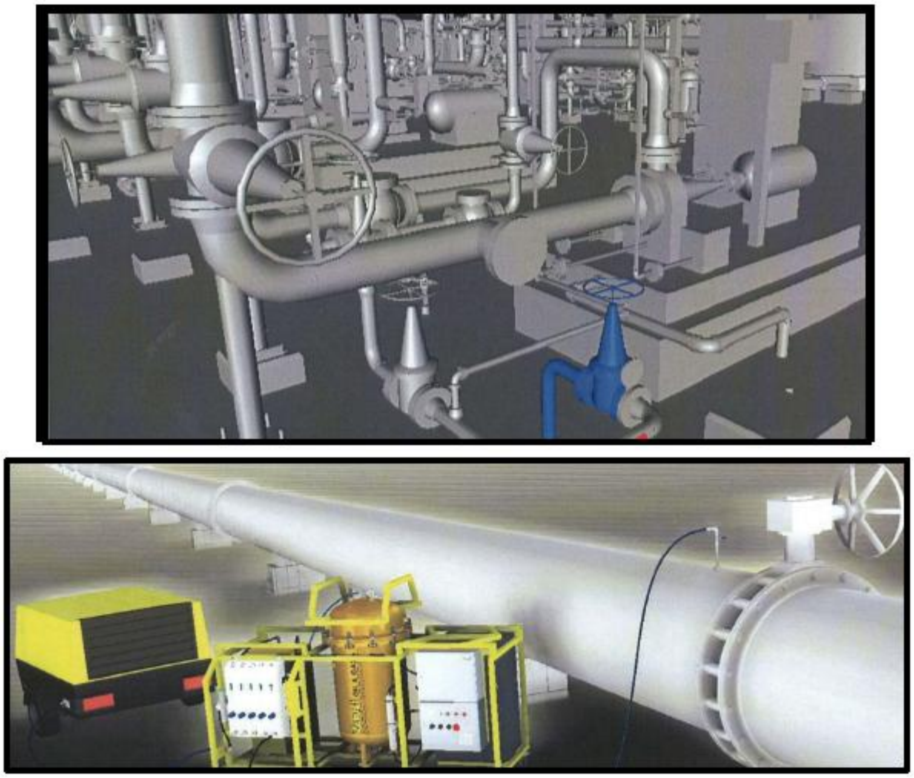


Figure 3.21: Mothaballing pipeline systems for applying DH-VCI (DVS-2) system (E. Lyublinski, (2015))

Chapter 4

Results & Discussion

4.1 Material and Energy Balance:

4.1.1 Introduction:

A material balance in it is most broad definition is the application of the law of conservation of mass, which states matter is neither created nor destroyed. Matter may flow through a control volume and may be reacted to form another species, however, no matter is ever lost or gained. The same is true for energy. As with material balances, we can apply the law that energy is neither created nor destroyed, it is simply converted into another form of energy. The law of conservation of mass and energy leads to what is called a mass (material) and energy balance. The data use in this chapter obtains from Aspen hysys model shown in appendix (B) *(A. Bhatia B.E. (2012)).*

4.1.2 Material balance:

4.1.2.1 Mass balance around Flash tower:

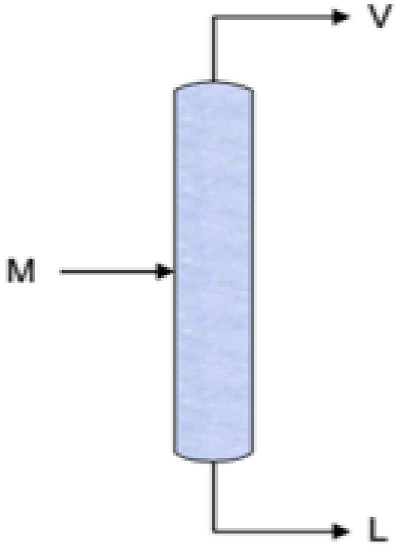


Figure 4.1: Flash tower

Total low = 311.159ton/h

V/M= (0.7545/760.2) = 0.0009925

At top:

V = 311.159*0.0009925 = 0.308825 ton/hr

At Bottom:

B = (1-0.0009925)*311.159 = 310.85 ton/hr

4.1.2.2 Material balance around Vacuum column:

The equation of material balance for any system=

Input + Generation – Consumption – Output = Accumulation

The mass flow rate:

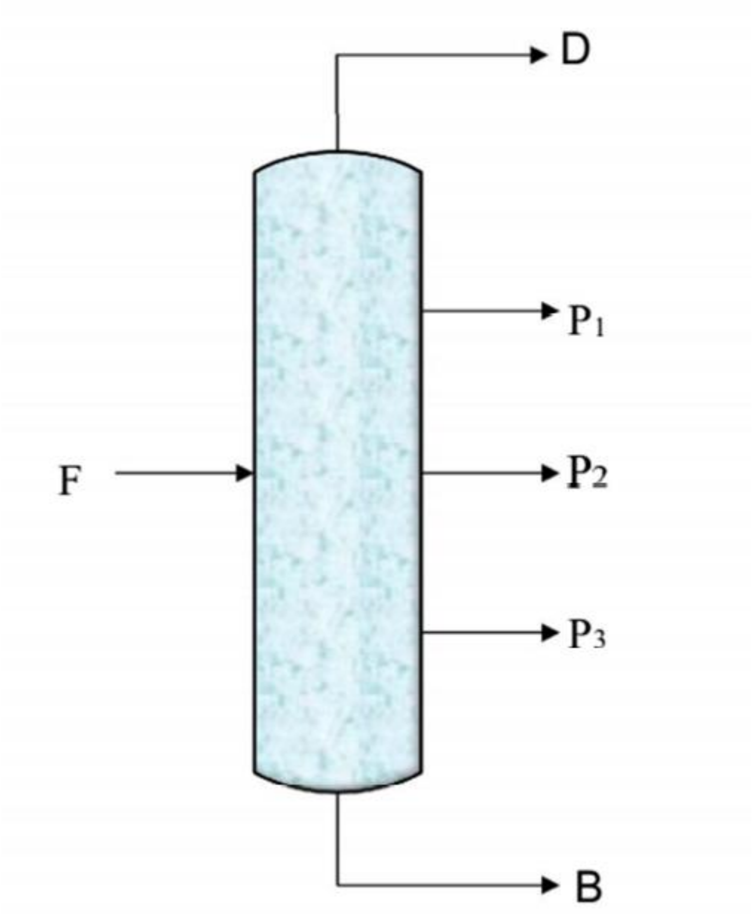


Figure 4.2: Vacuum column

At steady state:

- Accumulation=0
- Generation=0

- Consumption=0

Then:

Input = Output

For overall material balance

$$\boldsymbol{F = D + P_1 + P_2 + P_3 + R}$$

Where:

D = vacuum over head

F = vacuum Feed

P_1 = LVGO

P_2= HVG

P_3= Slop wax

R = vacuum residue

Table 4.1: Fractions and specific gravity of the product

Component	Concentration in feed	SG
D	0.0347	**0.840**
$\boldsymbol{P_1}$	0.0954	**0.7668**
$\boldsymbol{P_2}$	0.4233	**0.750**
$\boldsymbol{P_3}$	0.0208	**0.740**
R	0.4258	**0.7948**
Total F	**1**	

The mass Flow rate of: $\mathrm{D} = 311.159 \times 0.0347 = 10.8\ \mathrm{ton/h}$

The mass Flow rate of: $P_1 = 311.159 \times 0.0954 = 29.68\ ton/h$

The mass Flow rate of: $P_2 = 311.159 \times 0.4233 = 131.71\ ton/h$

The mass Flow rate of: $P_3 = 311.159 \times 0.0208 = 6.47\ ton/h$

The mass Flow rate of: $R = 311.159 \times 0.4258 = 132.5\ ton/h$

Table 4.2: Mass balance around vacuum tower

Products	**Cut mass %**	**Mass flow (Ton/h)**	**Mass flow (Kg/h)**	**Volume flow (m^3/h)**	**Volume flow (bbl/day)**	**Density**
vacuum over head	3.47	10.8	10810	237400	35880000	840
LVGO	9.54	29.68	29670	38.7	5842	766.8
HVGO	42.33		131700	175.6	26510	750
Slop wax	2.08	6.47	6479	8.755	1322	740
vacuum residue	42.58	132.5	132500	166.9	25170	794.8
Total	100	311.159	311159	354500	53520000	762

4.1.3 Energy Balance:

4.1.3.1 STEAM HEAT FLOW CALCULATION:

The first of thermodynamics can be written as fallow

Input of Energy - Output Energy = Accumulation of energy in process

We can assume

Steady state: accumulation= Zero

Rate of energy input= rate of energy output

Rate in = $\sum f_{i\,,in}\left(h_{i\,,in} + pe_{i\,,in} + ke_{i\,,in}\right) + Q + W$

Rate out = $\sum f_{i\,,out}\left(h_{i\,,out} + pe_{i\,,out} + ke_{i\,,out}\right)$

Where:

f_i= flow rate of component i

h_i=specific enthalpy,

pe_i=specific potential energy,

ke_i=specific kinetic energy

W=mechanical work done on system

Q = heat input to system

Changes in Ke and Pe smaller than enthalpy

$$\sum f_{i\ ,in}\left(h_{i\ ,in} + pe_{i\ ,in} + ke_{i\ ,in}\right) + Q + W$$

f_i , h_i=product of flow rate and specific enthalpy=total rate of energy transport with component i

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

$$\boldsymbol{H_T = \int_{T_d}^{T} C_p\, dT}$$

Where:

H_T= specific enthalpy at temperature T

C_p = specific heat capacity of the material, constant pressure,

T_d = the datum temperature

Table 4.3: Steam heat flow

	Mw	Mass-flow	Molar Enthalpy	Mass enthalpy	Heat flow
ATM	424.3	3.012e005	-5.851e005	-1379	**-4.15e008**
STEAM	18.02	907.2	-2.359e005	-1.309e004	**-1.188e007**
Vac- Feed	397.4	3.021e005	-4.484e005	-1128	**-3.40e008**
Vac-steam	18.02	9072	-2.342e005	-1.3e004	**-1.17e008**
Vac-overhead	19.02	1.069e004	-2.409e005	-1.266e004	**-1.354e008**
LVGO	260.3	2.266e004	-4.737e005	-1820	**-5.39e007**
HVGO	377.1	1.317e005	-5.785e005	-1534	**-2.02e008**
Vac-residue	590.9	1.326e005	-8.230e005	-1393	**-1.84e008**
Slop wax	467.7	6479	-6.312e005	-1350	**-8.744e006**
Heated-vac	**397.4**	**3.021e005**	**-5.620e005**	**-1414**	**-4.272e008**

4.1.3.2 Unit operation energy analysis:

Rate energy input = rate energy output

$$\boldsymbol{Q_{mass,in} = Q_{mass,out}}$$

$$\boldsymbol{Q = mC_p \Delta T}$$

Q = heat quantity or duty in Kj/hr

M = mass flow rate in kg/hr

C_p= specific heat capacity kj/kg.C°

ΔT = Temperature change in C°

T_o = Reference temperature 25 C°

4.1.3.2.1 Heater:

$3.021 \times 10^5 kg/hr$ of feed to be heated from (352.5 To 404.4)C° .

We assume steady state: accumulation=zero

Rate energy input=Rare energy output

$$\boldsymbol{m_1 C_{p_1} \Delta T_1 + Q_{in} = m_2 C_{p_2} \Delta T_2}$$

$$\boldsymbol{m_1 C_{p_1} \Delta T_1 + Q_{in}} = 3.021 \times 10^5 \times 1160\ (352.5 - 25) + 8.635 \times 10^7$$

$$= 1.149 \times 10^{11} kj/hr$$

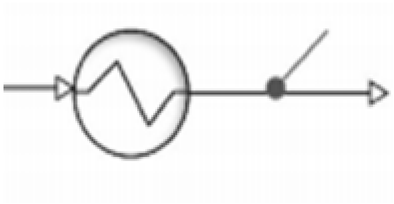
$$\boldsymbol{m_2 C_{p_2} \Delta T_2} = 3.021 \times 10^5 \times 1146\ (404.4 - 25) = 1.314 \times 10^{11}\ kj/hr$$

4.1.3.2.2 Pumps around energy balance:

Table 4.4: energy balance of pump around

	inlet T_1	Outlet T_2	Mass flow
Pump1(HVGO)	183.2C	21.23C	**132300**
Pump2(LVGO)	**294.5C**	**194.5C**	**313500**

4.1.3.2.2.1 Pumparound (HVGO):



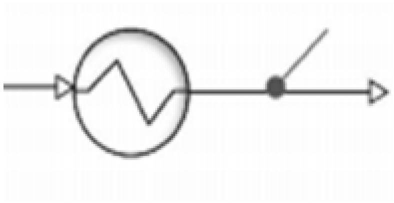
$$\boldsymbol{m_1C_{p_1}\Delta T_1 = Q + m_2C_{p_2}\Delta T_2}$$

$$\boldsymbol{m_1C_{p_1}\Delta T_1} = 132300 \times 1043 \times (183.2\text{-}25) = 2.183e + 010$$

$$\boldsymbol{Q + m_2C_{p_2}\Delta T_2} = 6.851\text{e}{+}007 + (3.135\text{e}{+}005 \times 934.4 \times (211.3 - 25)$$

$$= 5.464 \times 10^{6} kj/hr$$

4.1.3.2.2.2 Pumparound (LVGO):



$$\boldsymbol{m_1C_{p_1}\Delta T_1 = Q + m_2C_{p_2}\Delta T_2}$$

$$\boldsymbol{m_1C_{p_1}\Delta T_1} = 1.323e005 \times 637 \times (189.6 - 25) = 1.387 \times 10^{10} kj/hr$$

$$\boldsymbol{Q + m_2C_{p_2}\Delta T_2} = 5.663e + 007 + (1.323e + 005) \times 405.3 \times (-23.59 - 25\ )$$

$$= 1.387 \times 10^{10} kj/hr$$

Table 4.5: Unit operation duty

unit	Duty(kj/hr)
Heater	8.63E+007
Pump around 1	5.66E+007
Pump around 2	6.851E+007

4.2 Distillation design:

4.2.1 Introduction:

The separation of liquid mixtures into their various components is one of the major operations in the process industries, and distillation, the most widely used method of achieving this end, is the key operation in any oil refinery , that uses the difference in relative volatilities or differences in boiling of the component to be separated .

4.2.1.1 Types of distillation column:

1. Single flash vaporization
2. Packed towers
3. Plates tower
    a. Bubble cap towers
    b. Sieve pates
    c. Valve plates towers

4.2.1.2 General design methods:

In designing a column for a given separation, the number of stages required and the Flow rates of the liquid and vapour streams must first be determined using the general methods outlined previously. In the mechanical design of the column, tower diameter, tray spacing, and the detailed layout of each tray is considered. Initially, a diameter is established, based on the criterion of absence from liquid entrainment in the vapour stream, and then the weirs and the downcomers are designed to handle the required liquid flow.

4.2.1.3 Bubble cap trays:

Bubble-cap trays are rarely used for new installations on account of their high cost and their high pressure drop. In addition, difficulties arise in large columns because of the large hydraulic gradients which are set up across the trays. Bubble cap trays are capable of dealing with very low liquid rates and are therefore useful for operation at low reflux ratios. There are still many bubble-cap columns in use.

4.2.1.4 Sieve trays:

Sieve trays offer several advantages over bubble-cap trays, and their simpler and cheaper construction has led to their increasing use. The general form of the flow on a sieve tray is typical of a cross-flow system with perforations in the tray taking the place of the more complex bubble caps. the key differences in operation between these two types of tray should be noted. With the sieve tray the vapour passes vertically through the holes into the liquid on the tray, whereas with the bubble cap the vapour issues in an approximately horizontal direction from the slots. With the sieve plate the vapour velocity through the perforations must be greater than a certain minimum value in order to prevent the weeping of the liquid stream down through the holes. at the other extreme, a very high vapour velocity leads to excessive entrainment and loss of tray efficiency.

4.2.2 Heavy key and light key:

Heavy key (Hk):NBP [0]378*

Light key (Lk): NBP [0]302*

4.2.3 Type of tray:

Sieve tray

4.2.4 Calculation of relative volatility:

$$\boldsymbol{\alpha_i = \frac{K_i}{K_{hk}}}$$

$$\boldsymbol{\alpha_{avg} = \sqrt[3]{\alpha_{top}\alpha_{bot}\alpha_{feed}}}$$

Where :

K_i = K valu of component i

K_{hk} = K valu of heavy component

α from table 1, appendix

(calculation are made by using Excel sheet).

4.2.5 Determination of minimum Reflux ratio R_m :

$$\boldsymbol{\sum\frac{\alpha x}{\alpha-\theta} = R_m + 1} \; (1) \qquad \longrightarrow$$

Where:

$\boldsymbol{\alpha}$ = average Relative volatility of any component.

x = mole fraction of component.

$\boldsymbol{\theta}$ = constant.

$\boldsymbol{R_m}$= minimum Reflux ratio.

$$\sum \frac{\alpha Z_f}{\alpha - \theta} = 1 - q \xrightarrow{\hspace{3cm}} \textbf{(2)}$$

Where :

$\boldsymbol{Z_f}$ **=**mole fraction of component in feed.

q = Feed quality.

$$q = \frac{H_G - H_F}{H_G - H_L}$$

Where :

$\boldsymbol{H_G}$ **=** Enthalpy of gas at the feed (KJ/Kmol)

$\boldsymbol{H_F}$ **=** Enthalpy of liquid at the feed (KJ/Kmol)

$\boldsymbol{H_L}$ **=** Enthalpy of feed (KJ/Kmol)

$$q = \frac{-302841{,}81 - (-420628{,}76)}{-302841{,}81 - (-759194{,}42)} = 0.2581$$

Substitute in equation (2) to find (θ)

$$\sum \frac{\alpha Z_f}{\alpha - \theta} = 1 - 0.2581$$

$$\sum \frac{\alpha Z_f}{\alpha - \theta} = 0.7419$$

from table 1, appendix (A)

Solving (θ) by Goal seek

$\theta = 2.9$

Substitute in equation (1) to find (R_m)

$R_m = 0.906556$

from table 1, appendix (A)

Solving by Goal seek

4.2.6 Calculation of Actual Reflux Ratio (R):

The rule of thumb is:

$$\mathbf{R = (1.2 \text{ ------- } 1.5)}\ \boldsymbol{R_m}$$

R = 1.2 R_m

R = 1.2*0.906556

R = 1.0879

4.2.7 Calculation of Minimum number of stagesN_{min}:

$$\boldsymbol{N_{min}} = \frac{\ln\left[\left(\frac{x_{lk}}{x_{hk}}\right)_D \times \left(\frac{x_{hk}}{x_{lk}}\right)_B\right]}{\ln \alpha_{lk}}$$

Where :

$\boldsymbol{X_{lk}}$= mole fraction of light key.

$\boldsymbol{X_{hk}}$= mole fraction of heavy key.

$\boldsymbol{\alpha_{lk}}$= average relative volatility of light key.

$\boldsymbol{N_{min}}$ = 6.6348 stages

4.2.8 Calculation of theoretical number of stages:

$$\frac{N-N_{min}}{N+1} = \mathbf{0.75}\left[\mathbf{1} - \left(R - R_{min}/R+1\right)^{\mathbf{0.566}}\right]$$

$$\frac{R-R_{min}}{R+1} = \frac{1.0879 - 0.9066}{1.0879 + 1} = 0.0868$$

$$\frac{N-N_{min}}{N+1} = 0.75\ [1 - (0.0868)^{0.566}\ ]$$

$$\frac{N-N_{min}}{N+1} = 0.5619$$

N = 16.421 stages

Solving by Goal seek

4.2.9 Calculation of the column efficiency (E$_\circ$):

$$\text{E}_\circ = \mathbf{51 - 32.5\,log}(\mu_a \alpha_a)$$

Where:

$\boldsymbol{\mu_a}$= the molar average liquid viscosity.

α_a= average relative volatility of the light key.

μ_a= 0.18

$E_\circ$ = 61.28 %

4.2.10 Calculation of actual number of stages (N_a):

$$\mathbf{N_a = \frac{N}{E_\circ}}$$

$$=\frac{16.42}{0.6128} = 26.795 \text{ stages}$$

4.2.11 Calculation of the height of the column (H_t):

$$\boldsymbol{H_t = N_a \times C + \frac{(N_a - 1)\times C}{10} + 0.2 \times H}$$

C = tray spacing

C = 0.6

$$0.8\ H_t = 27 \times 0.6 + \frac{(27-1)\times 0.6}{10}$$

$\boldsymbol{H_t}$ = 22.03 m

4.2.12 Calculation of the feed plate location:

$$\boldsymbol{\log\left[\frac{N_r}{N_s}\right] = 0.206\ \log\left[\left(\frac{B}{D}\right)\left(\frac{X_{f,hk}}{X_{f,lk}}\right)\left(\frac{X_{b,lk}}{X_{d,hk}}\right)^2\right]}$$

Where :

$\boldsymbol{N_r}$ = number of stages above the feed

$\boldsymbol{N_s}$= number of stages below the feed

B = molar flow bottom product

D = molar flow top product

$\boldsymbol{X_{f,hk}}$ = concentration of the heavy key in the feed

$\boldsymbol{X_{f,lk}}$ = concentration of the light key in the feed

$\boldsymbol{X_{b,lk}}$ = concentration of the light key if in the bottom product

$\boldsymbol{X_{d,hk}}$ = concentration of the heavy key in the top product

$$\log\left[\frac{N_r}{N_s}\right] = 0.206\ \log\left[\left(\frac{1379.593}{57.918}\right)\left(\frac{0.0278}{0.012}\right)\left(\frac{0.000156}{0.00006}\right)^2\right]$$

$$\log\left[\frac{N_r}{N_s}\right] = 0.523$$

$\boldsymbol{N_r}$ **= 17.4**

$\boldsymbol{N_s}$ **= 10.29**

The feed enter the column at tray no **10** from the bottom

4.2.13 Calculation of diameter of the column (D_c):

The following areas terms are use in the design:

A_c= total column cross sectional area

A_d= cross sectional area of down comer

A_n = net area available for vapour-liquid disengagement, normally equal to $A_c - A_d$

4.2.14 For a single pass plate:

A_a= active area, equal to $A_c - 2A_d$ for single-pass plates

A_h= hole area, the total area of all the active holes

A_p= perforated area (including blanked areas),

A_{ap}= the clearance area under the down comer apron

Top diameter calculation:

$$\boldsymbol{U_f = K\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}}$$

Where:

U_f = flooding vapor velocitym/s, based on the net column cross-sectional area A_n

K = constant obtained from figure (1) , appendix (A)

$$\boldsymbol{F_{LV} = \frac{L_w}{V_w}\sqrt{\frac{\rho_v}{\rho_l}}}$$

Where :

F_{LV} = The liquid-vapor flow factor in figure (1) appendix

L_w= liquid mass flow-rate, kg/s,

V_w = vapor mass flow-rate, kg/s.

Top diameter calculations:

$$\frac{L_w}{V_w} = \frac{R}{R+1} = \frac{1.0879}{1.0879+1} = 0.521$$

From ideal gas law:

$$\rho_v = \frac{T_o P M_{wt}}{T P_o V_o}$$

$$\rho_v = \frac{273.15 \times 6.867 \times 19.29}{339.06 \times 1 \times 22.4} = 4.764 \text{ Kg}/m^3$$

$$F_{LV} = 0.521 \times \sqrt{\frac{4.764}{783}} = 0.041$$

From figure (1) appendix (A)

K = 0.11

$$U_f = 0.11 \times \sqrt{\frac{783 - 4.764}{4.764}} = 1.406$$

Design velocity = 80% of U_f

U = 0.8 ×1.41= 1.1247 m/s

Net column area used in separation is

$$A_n = \frac{Q_v}{U}$$

Q_v = Volumetric flow rate of vapors

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

Mass vapor flow rate = 19670 kg/h

Vapor density= 4.764 kg/m3

$$Q_v = \frac{19670}{3600 \times 4.764}$$

Q_v= 1.1469m3/sec

Now, net area

$$A_n = \frac{Q_v}{U} = \frac{1.1469}{1.1247} = 1.0197 m^2$$

Assume that down comer occupies 12% of cross sectional Area (A_c) of column

$$A_d = 0.12 A_c$$

$$A_c = A_n + A_d$$

$A_n = A_c - 0.12A_c = 0.88A_c$

$A_c = \frac{A_n}{0.88} = \frac{1.1097}{0.88} = 1.1588 \ m^2$

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

$D_c = \left(\frac{4\ A_c}{\pi}\right)^{0.5} = 1.215$ m

$A_d = 0.12A_c$

$A_d = 0.12 \times 1.215 = 0.1458m^2$

4.3 Plate Design:

$A_a = A_c - 2A_d$

= 1.1588 – 2 (0.1458) = 0.8672 m^2

Hole area A_h take 10% A_a

$A_h = 0.1 \times A_a$

= 0.1 × 0.8672 = 0.0867 m^2

4.3.1 Weir length (L_w):

$A_d/A_c = 0.1458/1.1588 = 0.126$

From Figure (3) appendix

$\frac{\boldsymbol{L_w}}{\boldsymbol{D_d}} = 0.77$

$$D_d = \left(\frac{4A_d}{\pi}\right)^{0.5}$$

$$D_d = \left(\frac{4 \times 0.1458}{\pi}\right)^{0.5} = 0.431\ m$$

$L_w = 0.77 \times 0.431 = 0.3318\ m$

4.3.2 Perforated area (A_p):

From figure (4), appendix θ_c = 100˚

Angle substances at plate edge by imperforated strip = 180-100 =80˚

Hole size: 5 mm is the preferred size

Calming zone width = 50 mm

Mean Length imperforated

$= (0.431 - 0.05)\ \pi \times \left(\frac{80}{180}\right) = 0.5317$ m

Area of imperforated = 0.05 × 0.5317 = 0.0266 m^2

Mean length of calming zone

$= (0.431 - 0.05) \sin\left(\frac{100}{2}\right) = 0.2919\ m^2$

Area of calming zone = 2(0.2919 × 0.05) = 0.0292 m^2

Total area for perforations,

$A_p = 1.1588 - 0.0266 - 0.0292 = 1.103 m^2$

4.3.3 Determination of entrainment correlation (Ψ):

From figure (2), appendix (A)

At $F_{LV} = 0.041$ and 80% flooding

$\psi = 3.4 \times 10^{-3}$

$\varphi = \frac{\psi \times L}{1-\psi} = \frac{3.4*10^{-3} \times 10263.7}{1-3.4*10^{-3}} = 30.88$ kg/h

4.3.4 Weeping point:

Weeping will occur when $U_{omin} < U_{omin}$ calculated

$U_o = \frac{V}{\rho_v \times A_h} = \frac{1.967*10^4}{4.764 \times 0.0867 \times 3600} = 13.23$ m/s

Taking 70% turn down

$U_{omin} = 0.7 \times U_o = 0.7 \times 13.23 = 9.26$ m/s

U_{omin} Calculated $= \frac{K_2 - 0.9\,(25.4 - d_o)}{\rho_v{}^{0.5}}$

$d_o = 5$ mm

K_2 is function of $(h_w + h_{owmin})$

h_w = weir high = 23 mm

h_{owmin} Minimum weir crest $= 750 \left(\frac{L_{min}}{\rho_l l_w}\right)^{2/3}$

L_{min} = 0.7×10263.7= 7184.59 kg/h

$h_{owmin} = 750 \times \left(\frac{7184.59}{783 \times 0.3318 \times 3600}\right)^{2/3} = 29.20\ mm$

$h_{owmin} + h_w = 29.20 + 23 = 52.2\ mm$

From figure (5), appendix (A)

K_2 = 30

U_{omin} calculated $=\frac{30-0.9(25.4-0.005)}{(4.764)^{0.5}} = 1.32\ m/s$

$U_{omin} > U_{omin}$ Calculated

Weeping will not occur

4.3.5 Pressure drop calculation:

ΔP = $9.81 \times h_t \times 10^{-3} \times \rho_l$

ΔP = total plate pressure drop, Pa (N/m2),

h_t = total plate drop, mm liquid

$$h_t = h_d + (h_w + h_{ow}) + h_r$$

$$h_d = 51 \left(\frac{U_h}{C_o}\right)^2 \frac{\rho_v}{\rho_l}$$

From figure (6), appendix (A)

At $\frac{A_h}{A_p} = \frac{0.0867}{1.103} = 7.86\%$

Plate thickness = 50 mm

Hole diameter = $d_h = \left(\frac{4\ A_h}{\pi}\right)^{0.5} = \left(\frac{4 \times 0.0867}{\pi}\right)^{0.5} =$ 0.33 m

$$\frac{\text{Plate thicknees}}{\text{Hole diameter}} = \frac{0.05}{0.33} = 0.2$$

$$C_o = 0.7$$

$h_d = 51 \left(\frac{13.23}{0.7}\right)^2 \times \frac{4.764}{783}$ = 110.84 mm

$$h_r = \frac{12500}{\rho_l} = \frac{12500}{783} = 15.96\ mm$$

h_w = 23mm

h_{ow} = 29.20mm

$$h_t = 15.96 + 110.84 + 23 + 29.20 = 179\ mm$$

ΔP = $9.81 \times 179 \times 10^{-3} \times 783$ = 1.37 kpa

4.3.6 Down comer liquid back up:

For safe design and to avoid flooding

$$h_b < \frac{1}{2}(C + h_w)$$

C = tray spacing

$h_b = h_t + h_{ow} + h_w + h_{dc}$

h_b= down comer back-up, measured from plate surface, mm

h_{dc} = head loss in the down comer, mm

$$h_{dc} = 166\left(\frac{L}{\rho_L A_m}\right)^2$$

A_m = either the down comer area A_d or the clearance area under the downcomer

A_{ap}; whichever is the smaller,m^2.

L = liquid flow rate in down comer, kg/s

$A_m = A_{ap} = h_{ap} \times L_w$

$h_{ap} = h_w - 10\ mm$

$h_{ap} = 23 - 5 = 18\ mm$

$A_{ap} = 0.3318 \times 18 \times 10^{-3} = 0.006\ m^2$

$h_{dc} = 166\left(\frac{10263.7}{783 \times 0.006 \times 3600}\right)^2$ = 61.13 mm

$h_b = 179 + 29.20 + 23 + 61.13 = 292.33\ mm = 0.292\ m$

$\frac{1}{2}(C + h_w)$= 0.5 (0.6+(23*10^{-3}) = 0.312 m

$h_b < \frac{1}{2}(C + h_w)$no flooding will occure.

4.3.7Down comer residence time:

$$\mathbf{t_r} = \frac{\mathbf{A_d h_b \rho_l}}{\mathbf{L}}$$

Where:$\mathbf{t_r}$ = residence time, s,

$$\mathbf{t_r} = \frac{0.1458 \times 0.292 \times 783}{10263.7/3600} = 11.7\ \text{sec}$$

4.4 Cost Estimation:

4.4.1 Introduction:

A preliminary economic analysis is performed for the overall plan. Due to lack of recent data, different cost estimates are done based on cost indices and capacity. However, the present analysis will give a fair idea about the profitability of the plant. Since the exact cost of the plant is not found, the calculations are done based on the purchased equipment cost (PEC).

4.4.2 Estimation of total capital investment:

Total capital investment (T c I) = fixed capital investment + working capital

Fixed capital investment= direct cost + indirect cost

4.4.2.1 Direct cost:

Material and labour involved in actual installation of complete facility (70-85% of fixed-capital investment) and we obtain this value from economic evaluation of ASPEN Hysys software this shown in table 6.1.The full sheet of project summary shown in Appendix (C).

4.4.2.2 Indirect costs:

Expenses which are not directly involved with material and labor of actual installation of complete facility (15-30% of Fixed-capital investment).

- Engineering and Supervision: (5-30% of direct costs)

Consider the cost of engineering and supervision = 10% of Direct costs

cost of engineering and supervision = 10% of 6.34E+07

- Construction Expense and Contractor’s fee: (6-30% of direct costs)

Consider the construction expense and contractor’s fee = 15% of Direct costs

i.e., construction expense and contractor’s fee = 15% of 6.34E+07

- Contingency :(5-15% of Fixed-capital investment or 20% to 40% of PEC)

Consider the contingency cost = 30% Purchased Equipment Cost

Contingency cost = 30% of 2.56E+07

Thus, Fixed capital investment = direct cost + indirect cost

= 6.34E+07 + 24.49E+06

= 8.69E+07$

4.4.2.3 Working Capital:

Consider the Working Capital = 15% of Fixed-capital investment

Working capital = 15% of FCI

= 13.04E+07$

Thus, Total capital investment = 8.69E+07 + 13.04E+07

= 21.73E+07$

4.4.3 Estimation of Total Product cost:

4.4.3.1 Manufacturing Cost:

Manufacturing cost = Direct production cost + Fixed charges + Plant overhead cost.

4.4.3.2 Fixed Charges:

Depreciation: (depends on life period, salvage value and method of calculation-about 10% of FCI for machinery and equipment and 2-3% for Building Value for Buildings)

Depreciation = (0.10×8.69E+07) + (0.02×38590000)

= 9451800$

- Local Taxes: (1-4% of fixed capital investment)

Consider the local taxes = 3% of fixed capital investment

= 0.3×8.69E+07

=26070000$

- Insurances: (0.4-1% of fixed capital investment)

Consider the Insurance = 1.0% of fixed capital investment

Insurance = 0.01×8.69E+07

=869000$

Thus, Fixed Charges= 36390800$

4.4.3.3 Direct Production Cost:

-Raw Materials: (10-50% of total product cost)

Consider the cost of raw materials = 30% of total product cost

Raw material cost = 30% of X

- Operating Labor (OL): (10-20% of total product cost)

Consider the cost of operating labor = 15% of total product cost

operating labor cost = 15% of X

- Direct Supervisory and Clerical Labor (DS & CL): (10-25% of OL)

Consider the cost for Direct supervisory and clerical labor = 20% of OL

Direct supervisory and clerical labor cost = 20% of 0.15X

= 0.2×0.15X

= 0.03X

- Utilities: (10-20% of total product cost)

Consider the cost of Utilities = 15% of total product cost

Utilities cost= 15% of X

- Maintenance and repairs (M & R): (2-10% of fixed capital investment)

Consider the maintenance and repair cost = 5% of fixed capital investment

Maintenance and repair cost = 0.05×8.69E+07

= 4345000$

-Operating Supplies: (10-20% of M & R or 0.5-1% of FCI)

Consider the cost of Operating supplies = 10% of M & R

= 0.10 × 4345000

= 434500$

-Laboratory Charges: (10-20% of Operating supplies)

Consider the Laboratory charges = 15% of Operating supplies

= 0.15 × 434500

= 65175$

-Patent and Royalties: (0-6% of total product cost)

Consider the cost of Patent and royalties = 5% of total product cost

Patent and Royalties = 5% of X

Thus, Direct Production Cost = 0.68X + 4844675

4.4.3.4Plant overhead Costs:

It is about5-15% of total product cost; includes for the following: general plant upkeep and overhead, payroll overhead, packaging, medical services, safety and protection, restaurants, recreation, salvage, laboratories, and storage facilities.

Consider the plant overhead cost = 10% of total product cost.

Thus, Manufacture cost = Direct production cost + Fixed charges + Plant overhead costs.

Manufacture cost = 0.68X + 4844675+ 36390800 + 0.1X

= 0.78X + 41235475

4.4.4 General Expenses:

General expenses **=** Administrative costs + distribution and selling costs +research and development costs

4.4.4.1 Administrative costs:

About 15% of costs for operating labor, supervision, and maintenance or 2-6% of total product cost. Includes costs for executive salaries, clerical wages, legal fees, office supplies, and communications.

Consider the Administrative costs = 4% of total product cost

4.4.4.2 Distribution and Selling costs:

It equal2-20% of total product cost includes costs for sales offices, salesmen, shipping, and advertising.

Consider the Distribution and selling costs = 10% of total product cost

4.4.4.3 Research and Development costs:

It is about 5% of total product cost. Consider the Research and development costs = 5% of total product cost.

Thus, General Expenses = 0.19 of total product cost

Thus, Total Product cost = Manufacture cost + General Expenses

$X = 0.78X + 41235475 + 0.19X$

$X = 13.75E08\$$

Table 4.6: Direct cost

Direct cost	Total Cost
Purchased Equipment Cost	2.56E+07
Equipment Setting	678996
Piping	9.75E+06
Civil	3.49E+06
Steel	410240
Instrumentation	1.45E+06
Electrical	523548
Insulation	2.38E+06
Paint	105782
Other	1.73E+07

G and A Overheads	1.74E+06
Total Cost	6.34E+07$

4.5 Result:

4.5.1 Introduction:

In this chapter we mention all result from each section.

4.5.2 Column profiles:

- Pressure profile:

Figure 4.3 shown the increasing of pressure during process.

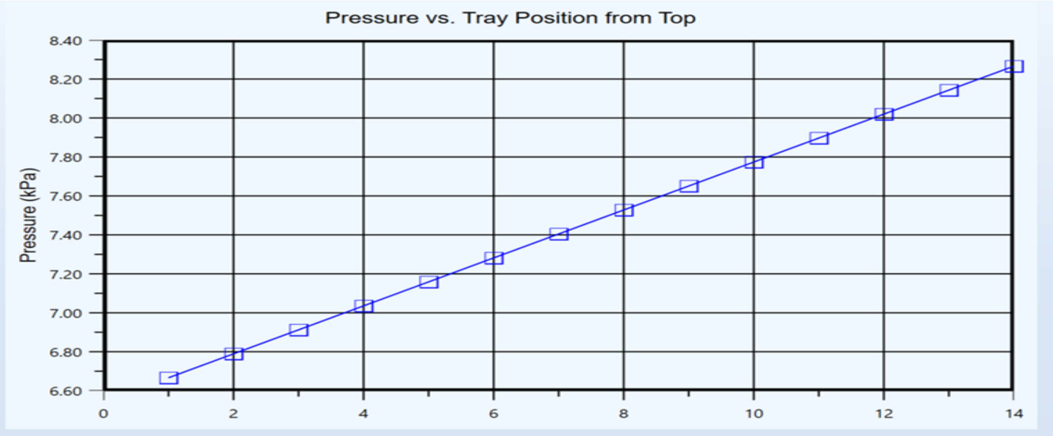


Figure 4.3: pressure profile

- Temperature profile:

The change of temperature from tray to another shown in figure 4.4

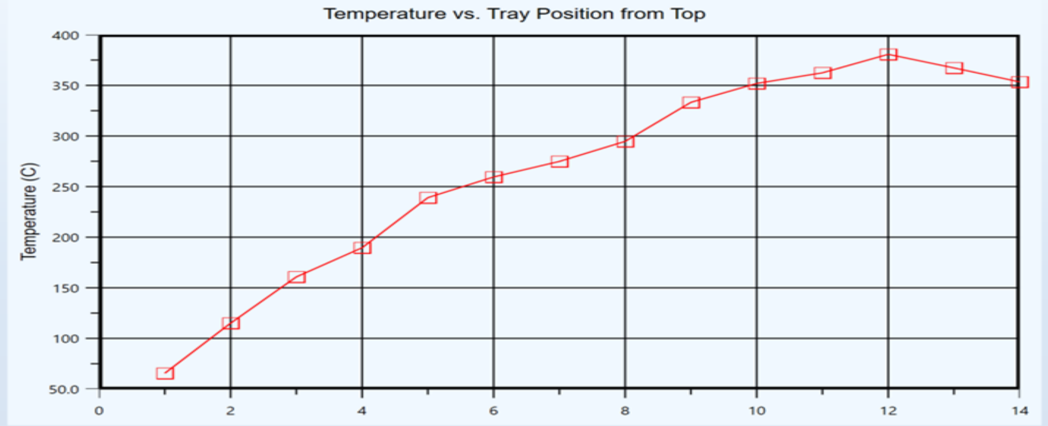


Figure4.4: temperature profile

- Net flow profile:

The quantity of liquid flow and vapour flow shown in figure 4.5.

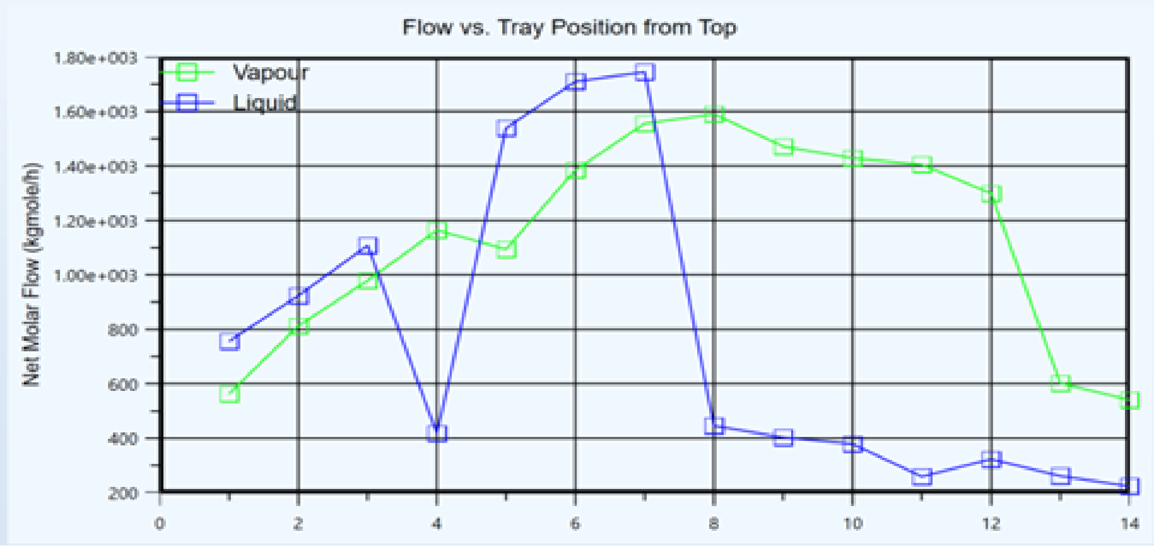


Figure 4.5: Net flow profile

- Transport properties:

The change of viscosity, density and molecular weight through trays shown in figure 4.6 and figure 4.7.

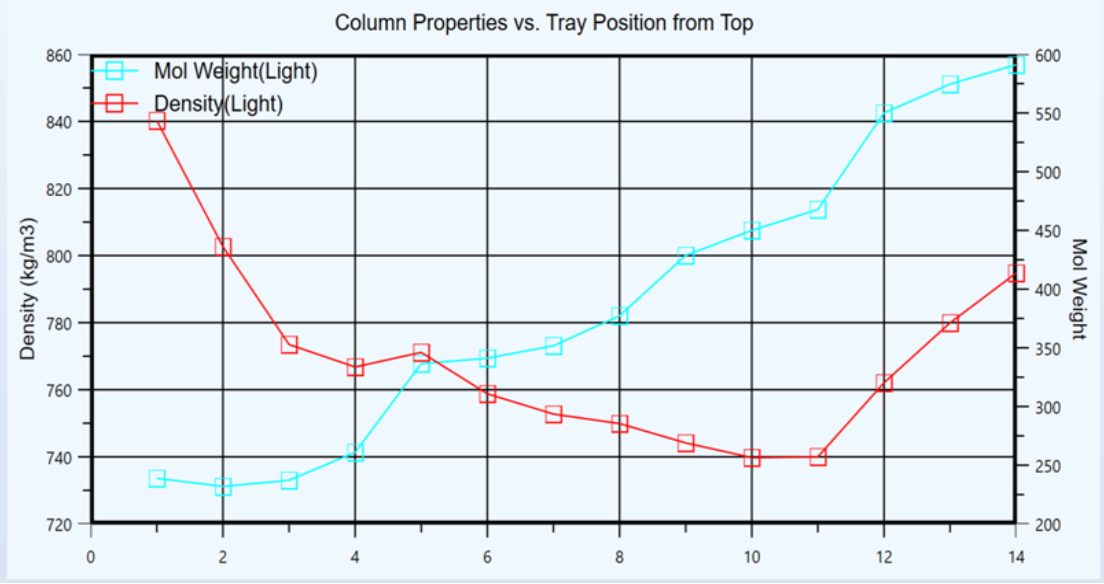


Figure 4.6: Density profile

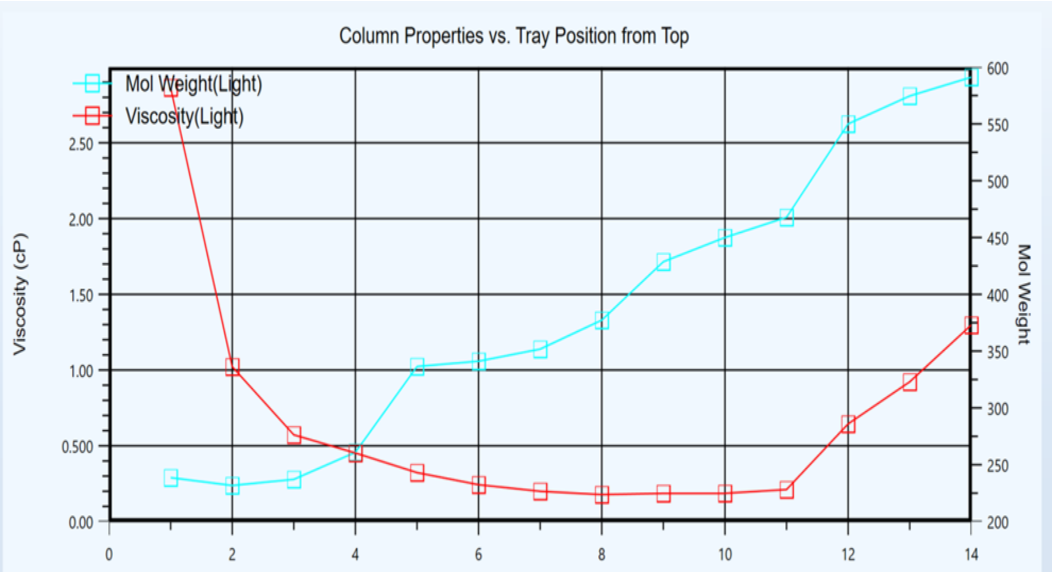


Figure 4.7: Viscosity profile

4.5.3 Improving of production result:

The result from use specs summary from ASPEN Hysys software that we previously explains shown in table 4.7 before increasing and after increasing shown in table 4.8 for light vacuum gas oil (LVGO) and for heavy vacuum gas oil (HVGO) shown in table 4.9 before increasing and after increasing show in table 4.10.The flow rate chart after improving the production of LVGO and HVGO is shown in

Table 4.7: before improving LVGO

Flow rate	**29673.11kg/hr**
Density @15 ºC	764.8 kg/m3
Molecular weight	254.6
Kinematic viscosity	0.5796 (cSt)

Table 4.8: After improving LVGO

Flow rate	**112489.26 kg/hr**
Density @15 ºC	783.9 kg/m3
Molecular weight	325.7
Kinematic viscosity	0.5324 (cSt)

Table 4.9: before improving HVGO

Flow rate	**131678.155 kg/hr**
Density @15 ºC	757.5kg/m3
Molecular weight	405.3
Kinematic viscosity	0.3223 (cSt)

Table 4.10: After improving HVGO

Flow rate	**184171.493kg/hr**
Density @15 ºC	757.5kg/m3
Molecular weight	405.3
Kinematic viscosity	0.3223 (cSt)

4.5.4 Design results:

The result from theoretical design of vacuum column shown in table 4.11.

Table 4.11: design result

No. of tray	27 tray	Perforated area	1.103 m²
Pressure drop	1.37 kpa	Tray thickness	0.05 m
Height of column	22.03 m	Weir height	0.023 m
Diameter of column	1.2 m	Reflux ratio	1.09
Hole size	0.05 m	Tray spacing	0.6
Weir length	0.3318 m	Active area	1.159 m

4.5.5 Cost estimation:

Cost estimation results summarize in a table 4.12, table 4.13, table 4.14 and table 4.15.

Table 12.6: Fixed capital cost

PROJECT CAPITAL SUMMARY	Total Cost
direct cost	6.34E+07
Indirect costs	24.49E+06
Fixed capital investment	8.69E+07

Table 4.13: Total capital investment

PROJECT CAPITAL SUMMARY	Total Cost
Fixed capital investment	8.69E+07
Working capital	13.04E+07
Total capital investment	21.73E+07

Table 4.14: Manufacture cost

PROJECT CAPITAL SUMMARY	Total Cost
Direct production cost	93.98EO7
Fixed charges	36390800
Plant overhead cost.	13.75E07
Manufacture cost	111.75E07

Table 4.15: Total product cost

PROJECT CAPITAL SUMMARY	Total Cost
General Expenses	26.125E07
Manufacture cost	111.75E07
Total Product cost	13.75E08

4.5.6 Aspen Hysys Result:

The full model of the process shown in figure 4.8.

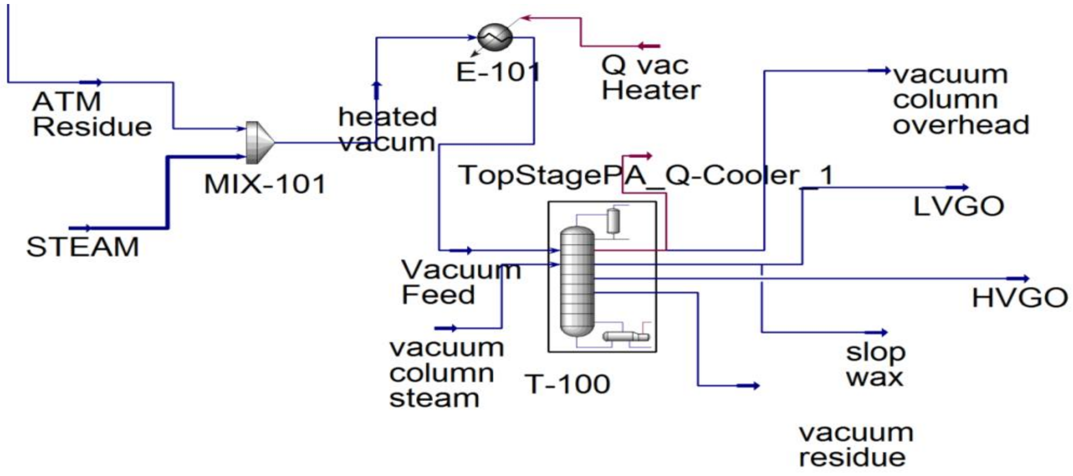


Figure 4.8: Overall view of vacuum distillation process.

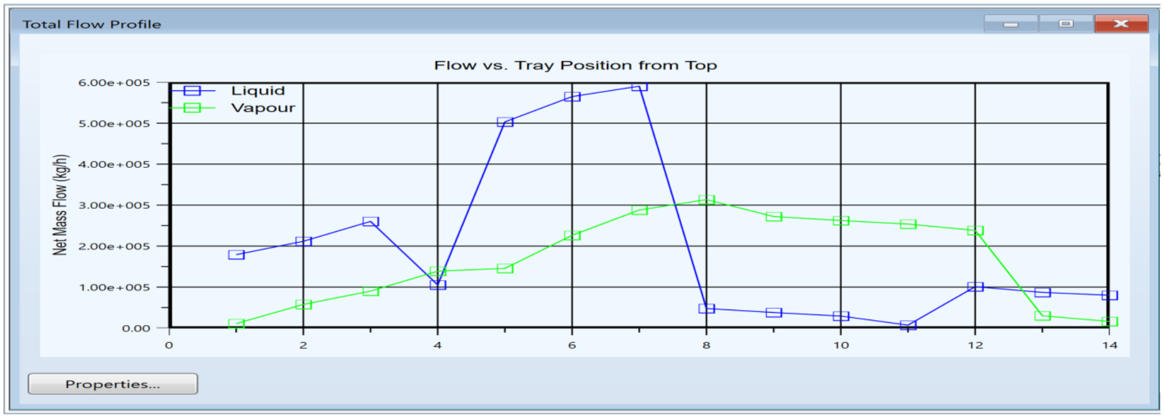


Figure 4.8: Net flow profile case of increasing HVGO

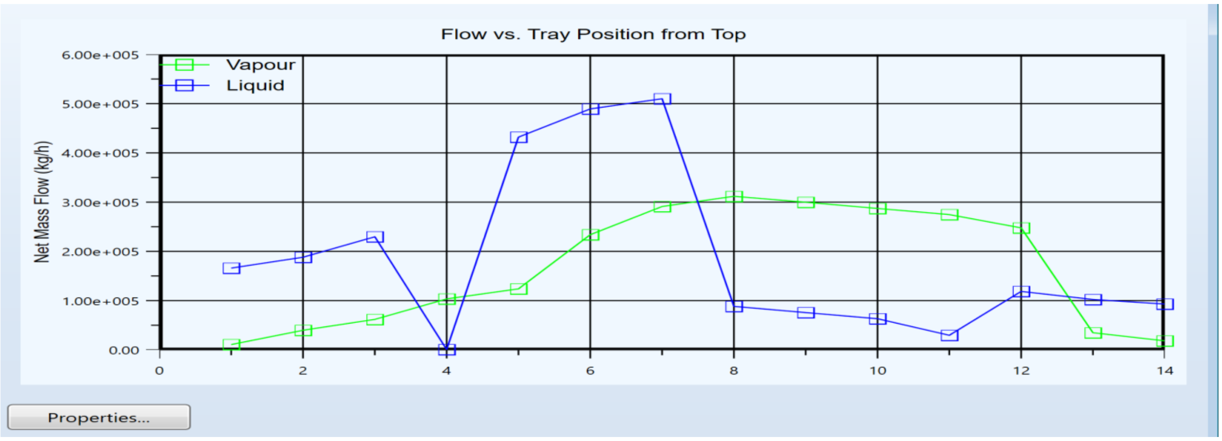


Figure 4.9: net flow profile case of increasing LVGO

Chapter 5

Conclusion

&Recommendation

5.1 Conclusion:

The designs have been performed using ASPEN Hysys software this represent a benefit and useful results. The material and energy balance have been calculated using information from ASPEN Hysys model this help to give full evaluate of the equipment. The theoretical design gives a specified details and information of design the vacuum distillation column. Cost estimation of process give a brief view of total cast that may help to anticipate the capital needed. By ASPEN Hysys software we performed a method to increasing the production of light vacuum gas oil (LVGO) and heavy vacuum gas oil (HVGO) this give us a marked increase in a flow rate of production.

5.2 Recommendation:

- We recommend using vacuum distillation unit in Khartoum Refinery because of economic impact of it.
- For other prospector we advise them to visit the Khartoum refinery to evaluate all factors that may affect on the process.

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Appendix (A)

Equipment Design Chart

Appendix (A)

Equipment design chart

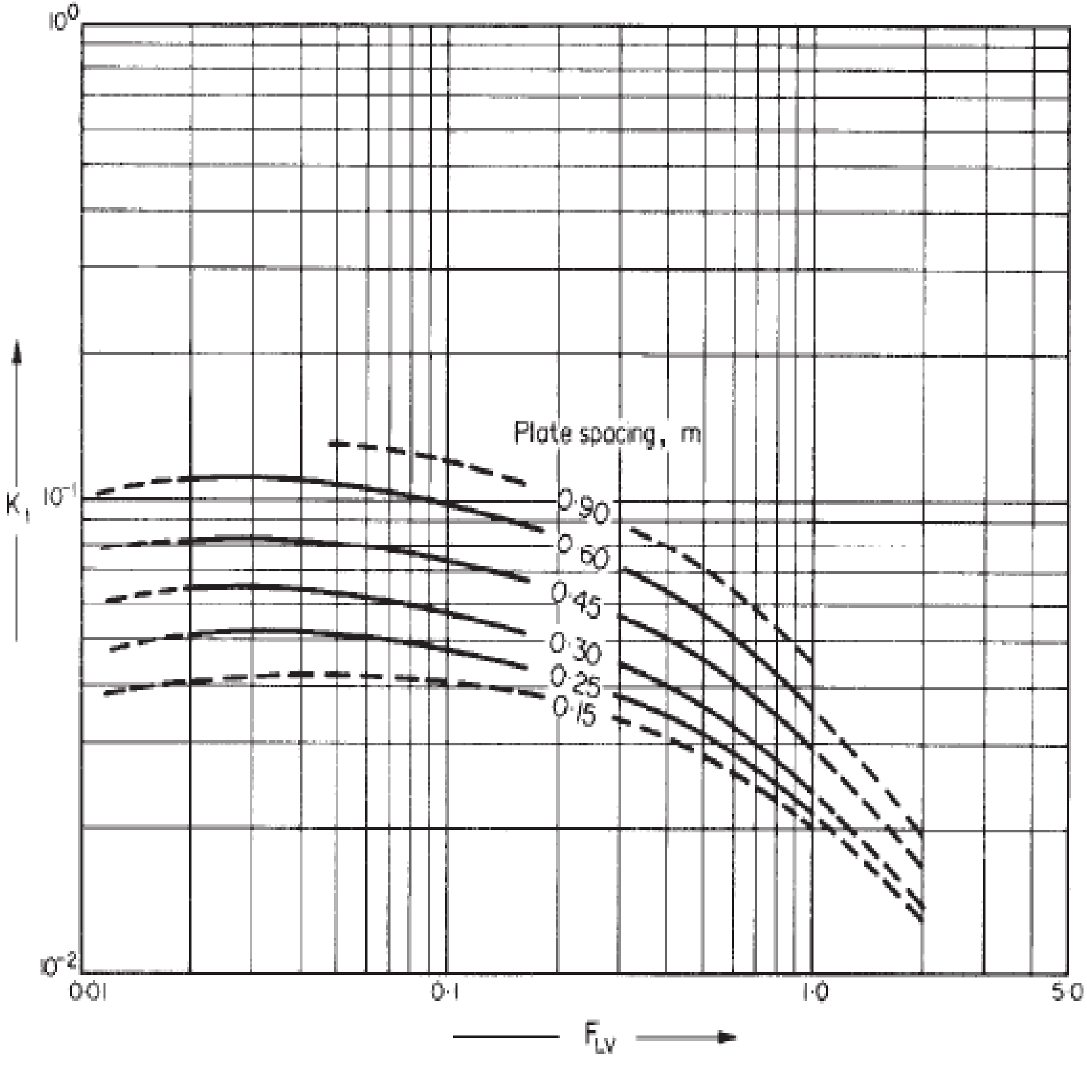

Figure 1: Flooding velocity, sieve plates

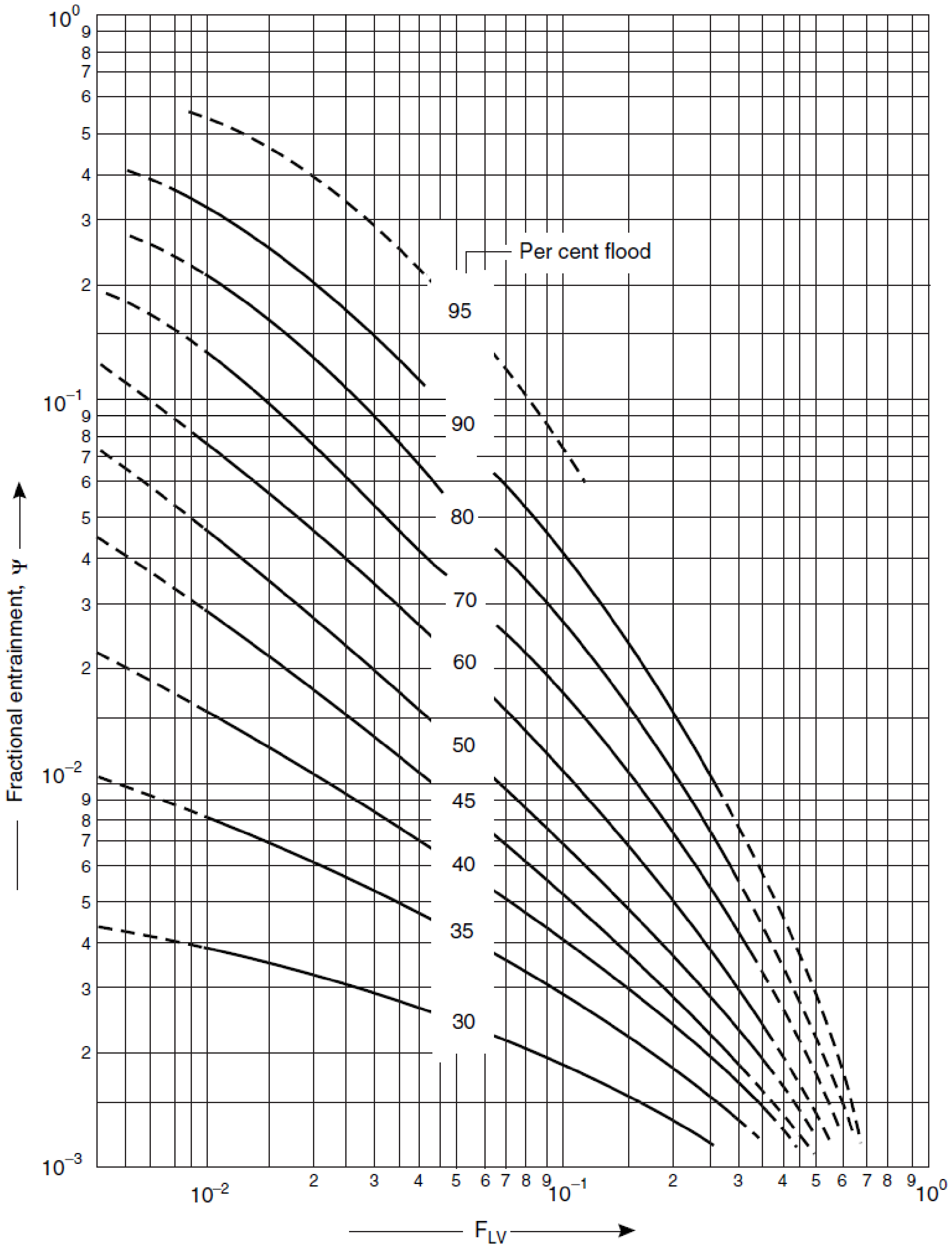

Figure 2: Entrainment correlation for sieve plates (Fair, 1961)

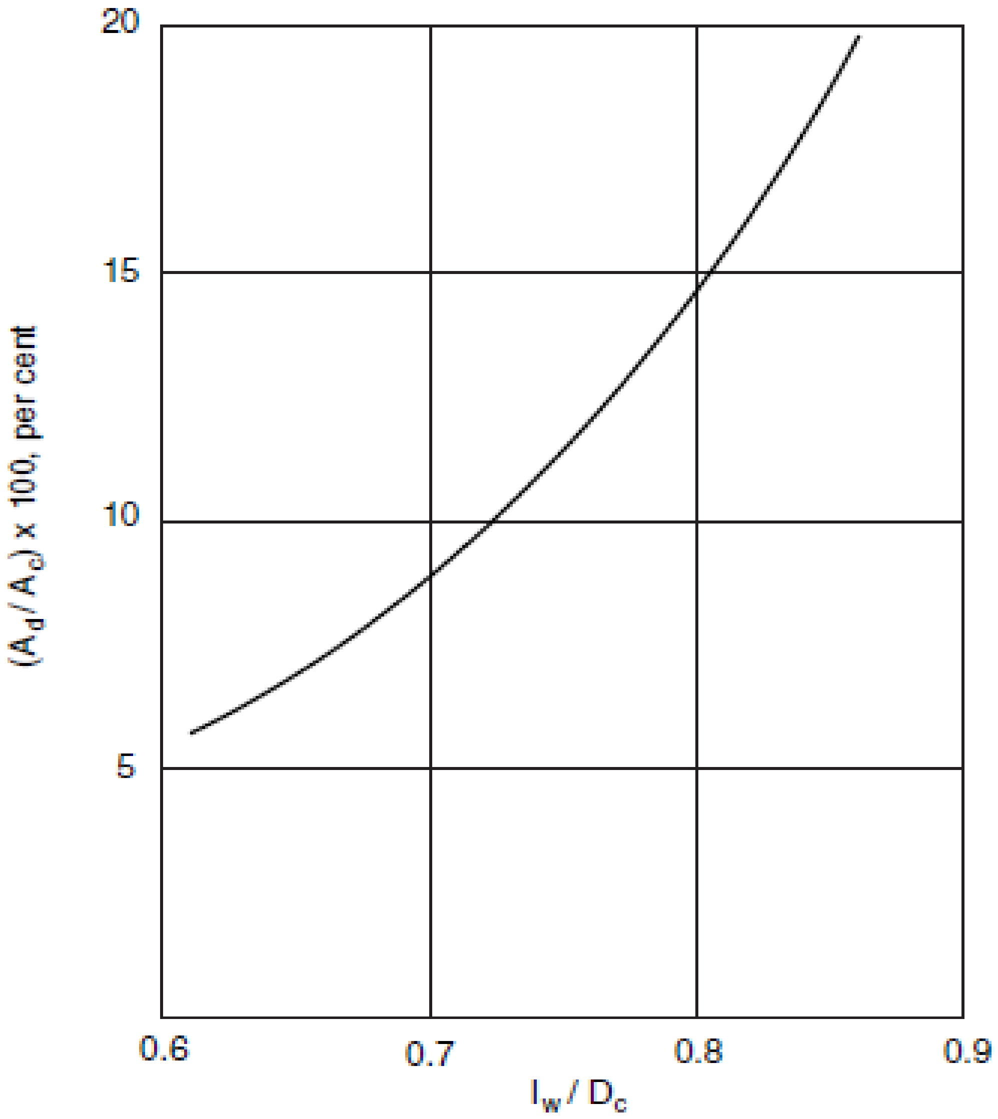

Figure 3: Relation between down comer area and weir length

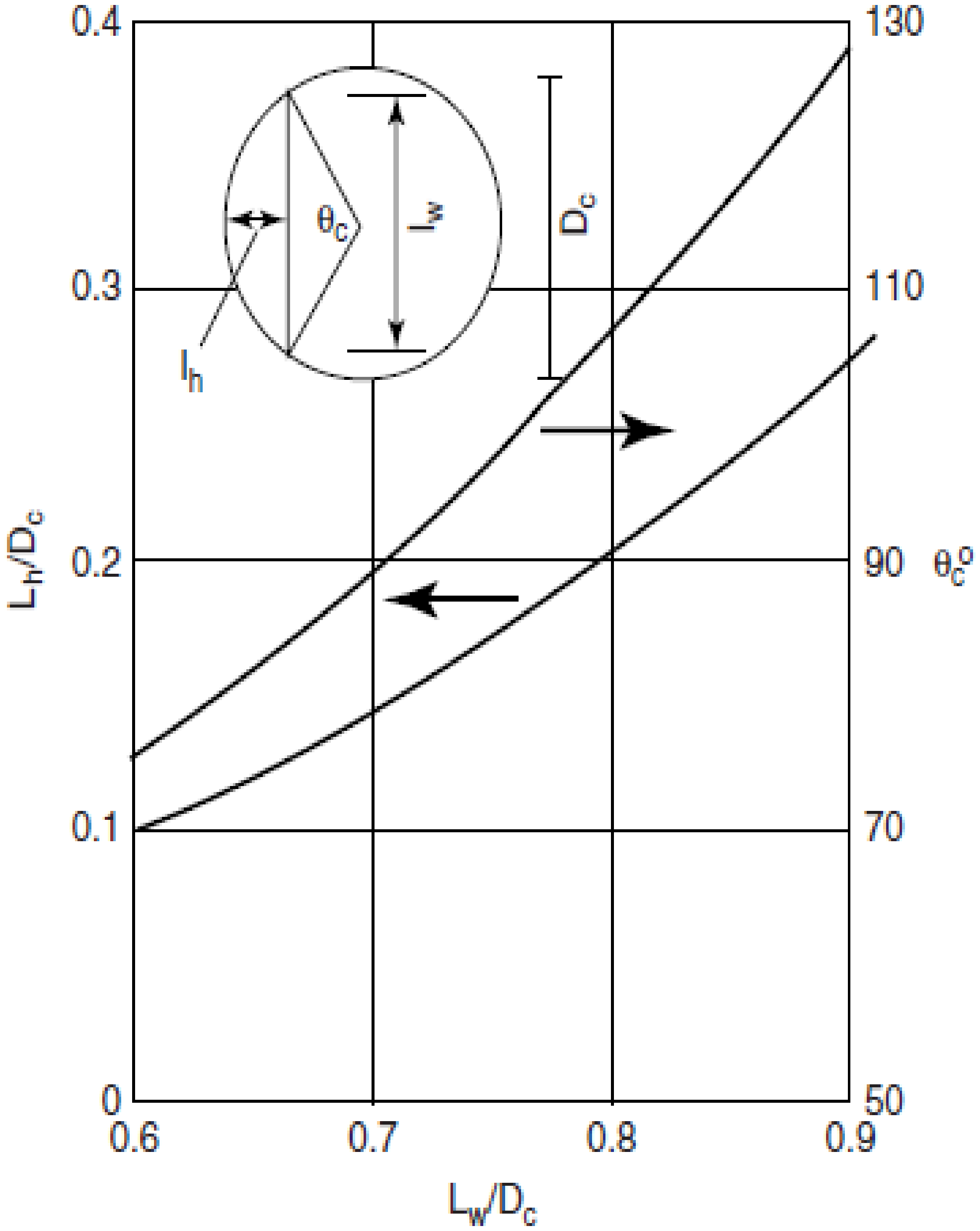

Figure 4: Relation between angle subtended by chord, chord height and chord length

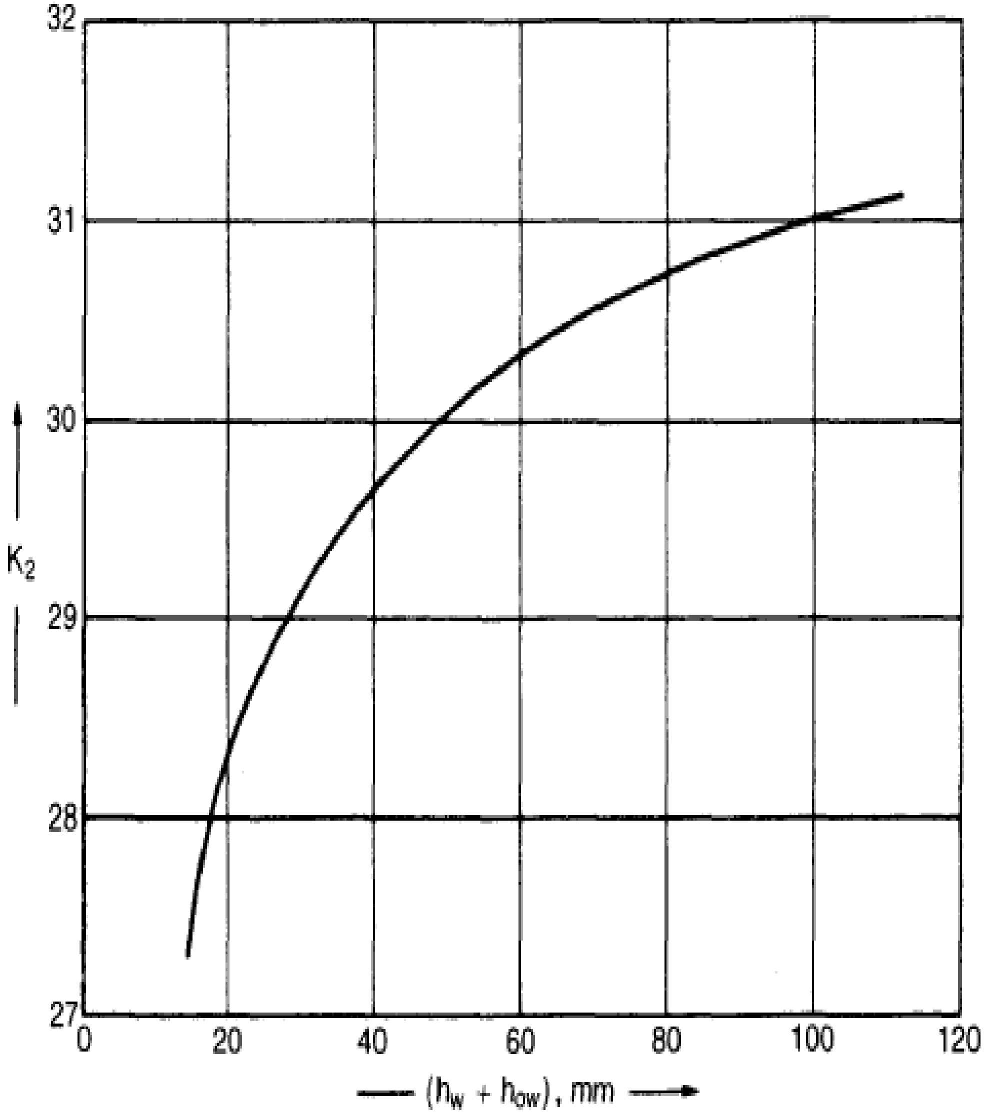

Figure 5: Weep-point correlation (Eduljee, 1959)

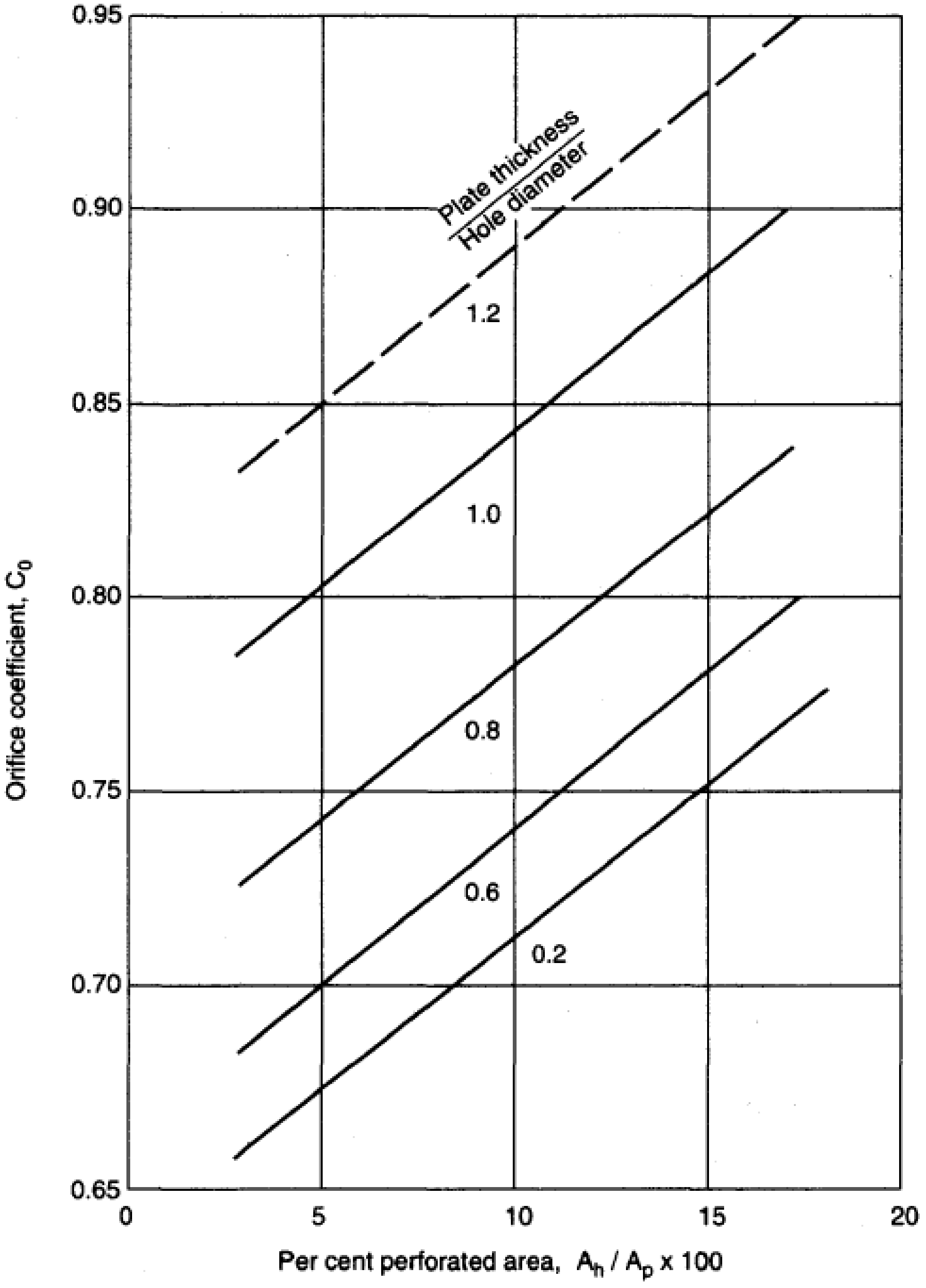

Figure 6:Discharge coefficient, sieve plates (Liebson*et al.*,1957)

Table 1: Design Calculation

Composition	K valu FEED	ALPH Feed	K valu top	ALPH Top	K valu Bot	ALPT Bot
Methane	0	0	0	0.00E+00	0	0
n-Pentane	973.373166	170.51079	236.3106257	1.84E+04	0	0
i-Pentane	1016.59624	178.08240	269.1185837	2.09E+04	0	0
n-Butane	1369.93278	239.97819	464.8793757	3.61E+04	0	0
i-Butane	1500.70404	262.88606	560.0789354	4.35E+04	0	0
Propane	0	0	0	0.00E+00	0	0
Ethane	0	0	0	0.00E+00	0	0
H2O	3057.74558	535.64105	2573.401861	2.00E+05	3018.0028	6029.686796
NBP[0]60*	694.935017	121.73535	138.4761383	1.08E+04	0	0
NBP[0]77*	574.282072	100.59995	97.5335138	7.58E+03	0	0
NBP[0]93*	473.218455	82.896115	68.31826135	5.31E+03	0	0
NBP[0]113*	376.822779	66.009988	44.06260123	3.42E+03	173.73041	347.0970929
NBP[0]135*	293.397484	51.395949	27.12838649	2.11E+03	122.79089	245.3246981
NBP[0]151*	239.273434	41.914761	18.33011603	1.42E+03	92.628397	185.0628591
NBP[0]170*	187.146263	32.783375	11.41046226	8.86E+02	65.895255	131.6525457
NBP[0]189*	144.988055	25.398305	6.978691321	5.42E+02	46.261051	92.4252463
NBP[0]208*	111.100974	19.462130	4.1789157	3.25E+02	31.977287	63.88762452
NBP[0]227*	83.9540552	14.706665	2.4382531	1.89E+02	21.680984	43.31657514
NBP[0]246*	63.3095612	11.090262	1.40319376	1.09E+02	14.600085	29.1696012
NBP[0]264*	47.1675087	8.2625756	0.79196764	6.15E+01	9.679122	19.337979
NBP[0]283*	34.5656831	6.0550488	0.432224024	3.36E+01	6.2650546	12.51699214
NBP[0]302*	24.9073259	4.3631446	0.228152583	1.77E+01	3.9587069	7.909125575
NBP[0]321*	17.7540677	3.1100715	0.11784609	9.15E+00	2.4623982	4.91964099
NBP[0]340*	12.3873166	2.1699501	5.83E-02	4.53E+00	1.4855727	2.968035016
NBP[0]359*	8.48310144	1.4860286	2.79E-02	2.16E+00	0.8729901	1.7441524
NBP[0]378*	5.70857215	1	1.29E-02	1.00E+00	0.500524	1
NBP[0]397*	3.7585863	0.6584109	5.71E-03	4.43E-01	0.2784022	0.556221602
NBP[0]416*	2.45971215	0.4308805	2.50E-03	1.94E-01	0.1535514	0.306781356
NBP[0]443*	1.24662324	0.2183774	6.69E-04	5.20E-02	5.92E-02	0.118183263
NBP[0]484*	0.42958316	0.0752523	0	0.00E+00	1.33E-02	0.026551507
NBP[0]521*	0.14791977	0.0259119	0	0.00E+00	3.00E-03	0.005985741
NBP[0]562*	4.02E-02	0.0070484	0	0.00E+00	4.89E-04	0.000975979
NBP[0]600*	1.09E-02	0.0019078	0	0.00E+00	8.00E-05	0.000159783
NBP[0]641*	2.40E-03	0.0004203	0	0.00E+00	9.97E-06	1.9919E-05
NBP[0]737*	4.84E-05	8.4773E-06	0	0.00E+00	5.00E-08	9.99721E-08

Table 1: Design Calculation

AVG ALPH	X D	X B	X F
0	5.36E-17	2.19E-24	4.05E-13
0	4.56E-10	4.80E-16	1.59E-07
0	2.35E-10	2.27E-16	9.35E-08
0	4.08E-11	2.11E-17	2.80E-08
0	2.87E-12	1.23E-18	2.38E-09
0	6.66E-13	1.58E-19	9.42E-10
0	3.67E-15	4.08E-22	1.07E-11
8642.987953	2.19E-04	2.88E-04	0.4189834
0	3.20E-09	6.83E-15	6.53E-07
0	1.33E-08	4.24E-14	1.91E-06
0	4.34E-08	2.06E-13	4.34E-06
428.0377668	1.32E-07	1.00E-12	8.45E-06
298.4030182	5.71E-07	7.23E-12	2.21E-05
222.7010812	2.21E-06	4.17E-11	5.62E-05
156.3989706	7.77E-06	2.23E-10	1.12E-04
108.3664752	3.89E-05	1.37E-09	2.45E-04
73.90300133	3.30E-04	8.91E-09	5.44E-04
49.41447615	4.72E-03	6.88E-08	1.36E-03
32.79204369	4.76E-02	4.84E-07	3.04E-03
21.42149732	0.17976056	2.89E-06	5.52E-03
13.65255745	0.30016649	1.65E-05	8.36E-03
8.488344402	0.28557563	1.56E-04	1.20E-02
5.193316903	0.1517678	2.35E-03	1.56E-02
3.079044062	2.82E-02	1.66E-02	1.87E-02
1.776980259	1.59E-03	3.54E-02	2.29E-02
1	6.24E-05	4.96E-02	2.78E-02
0.545558188	2.39E-06	6.80E-02	3.64E-02
0.295133816	9.31E-08	8.70E-02	4.57E-02
0.110289362	8.03E-10	0.1744296	9.06E-02
0	1.55E-13	0.1458417	7.53E-02
0	3.25E-17	0.120994	6.24E-02
0	1.04E-21	8.92E-02	4.60E-02
0	3.29E-26	6.10E-02	3.14E-02
0	2.14E-30	6.20E-02	3.20E-02
0	7.02E-30	8.72E-02	4.50E-02

Appendix (B)
Aspen Hysys Report

Column Sub-Flowsheet: T-100 @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION	
vacuum column steam	14__TS-1		
Vacuum Feed	12__TS-1	Heater	E-101

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
vacuum column overhead	1__TS-1	
vacuum residue	14__TS-1	
TopStagePA_Q-Cooler_1		
LVGO	4__TS-1	
HVGO	8__TS-1	
slop wax	11__TS-1	
HVGO pump around_Q-Cooler	HVGO pump around	
LVGO_Draw		
PA_1_Q-Cooler	PA_1	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error
LVGO pumparound_Rate(Pa)	22.30 kbpd *	22.30 kbpd	-4.206e-006
LVGO pumparound_Duty(Pa)	---	---	---
LVGO Rate	5.000 kbpd *	5.000 kbpd	-4.199e-006
HVGO Rate	21.00 kbpd *	21.00 kbpd	-7.073e-006
slop wax Rate	1.000 kbpd *	1.000 kbpd	-3.420e-005
LVGO pumparound_TRet(Pa)	85.00 F *	---	---
HVGO pump around_Rate(Pa)	50.00 kbpd *	50.00 kbpd	-7.061e-006
HVGO pump around_Dt(Pa)	150.0 F *	150.0 F	1.246e-006
TOP TEMPERATURE	150.0 F *	150.0 F	-1.995e-006
LVGO D1160 T95	915.0 F *	768.2 F	-1.631e-002
HVGO D1160 T95	1050 F *	955.1 F	-1.054e-002
NET FROM 4	0.1000 kbpd *	18.35 kbpd	182.5
NET FROM 11	3.000 kbpd *	26.80 kbpd	7.932
PA_1_Rate(Pa)	22.30 kbpd *	22.30 kbpd	-4.206e-006
PA_1_TRet(Pa)	85.00 F *	-10.47 F	-0.1061

	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
LVGO pumparound_Rate(Pa)	1.000e-002	22.19 kbpd *	On	On	On
LVGO pumparound_Duty(Pa)	1.000e-002	0.9478 Btu/hr	Off	On	Off
LVGO Rate	1.000e-002	22.30 kbpd *	On	On	On
HVGO Rate	1.000e-002	50.00 kbpd *	On	On	On
slop wax Rate	1.000e-002	0.9999 kbpd *	On	On	On
LVGO pumparound_TRet(Pa)	1.000e-002	1.800 F *	Off	On	Off
HVGO pump around_Rate(Pa)	1.000e-002	50.00 kbpd *	On	On	On
HVGO pump around_Dt(Pa)	1.000e-002	1.800 F *	On	On	On
TOP TEMPERATURE	1.000e-002	118.0 F *	On	On	On
LVGO D1160 T95	1.000e-002	0.1800 F	Off	On	Off
HVGO D1160 T95	1.000e-002	0.1800 F	Off	On	Off
NET FROM 4	1.000e-002	0.1510 kbpd	Off	On	Off
NET FROM 11	1.000e-002	0.1510 kbpd	Off	On	Off
PA_1_Rate(Pa)	1.000e-002	0.1510 kbpd	Off	On	Off
PA_1_TRet(Pa)	1.000e-002	1.800 F *	Off	On	Off

SPECS

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

Column Specification Parameters

LVGO pumparound_Rate(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	Flow Rate	Pumparound:	PA_1	Flow Basis:	Std Ideal Vol		

LVGO pumparound_Duty(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	Duty	Pumparound:					

LVGO Rate

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stream:	LVGO	Flow Basis:	Std Ideal Vol				

HVGO Rate

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stream:	HVGO	Flow Basis:	Std Ideal Vol				

slop wax Rate

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stream:	slop wax	Flow Basis:	Std Ideal Vol				

LVGO pumparound_TRet(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	eturn Temperature	Pumparound:					

HVGO pump around_Rate(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	Flow Rate	Pumparound:	VGO pump around	Flow Basis:	Std Ideal Vol		

HVGO pump around_Dt(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	Temperature Drop	Pumparound:	VGO pump around				

TOP TEMPERATURE

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stage:	1__TS-1						

LVGO D1160 T95

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stage:	4__TS-1	Type:	D1160 ATM	Flow Basis:	Volume Fraction	Phase:	Liquid
Cut Point	95.00 *						

HVGO D1160 T95

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stage:	7__TS-1	Type:	D1160 ATM	Flow Basis:	Volume Fraction	Phase:	Liquid
Cut Point	95.00 *						

NET FROM 4

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Stage:	4__TS-1	Flow Basis:	Std Ideal Vol	Liquid Spec:	Light		

NET FROM 11

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---

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

Column Specification Parameters

NET FROM 11

Stage:	8__TS-1	Flow Basis:	Std Ideal Vol	Liquid Spec:	Light		

PA_1_Rate(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	Flow Rate	Pumparound:	PA_1	Flow Basis:	Std Ideal Vol		

PA_1_TRet(Pa)

Fix/Rang:	Fixed	Prim/Alter:	Primary	Lower Bnd:	---	Upper Bnd:	---
Spec Type:	eturn Temperature	Pumparound:	PA_1				

PROFILES

General Parameters

Sub-Flow Sheet:	T-100 (COL2)	Number of Stages:	14 *

Profile Estimates

	Temperature (F)	Net Liquid (lbmole/hr)	Net Vapour (lbmole/hr)
1__TS-1	150.0 *	1667	1241
2__TS-1	325.0 *	2035	1790
3__TS-1	321.4	2442	2158
4__TS-1	373.2	920.1	2565
5__TS-1	462.3	3393	2412
6__TS-1	499.1	3769	3053
7__TS-1	526.7	3846	3429
8__TS-1	562.4	981.9	3506
9__TS-1	631.9	888.4	3243
10__TS-1	665.4	835.6	3150
11__TS-1	684.4	571.4	3097
12__TS-1	717.2	711.5	2863
13__TS-1	693.0	576.4	1328
14__TS-1	700.0 *	494.2	1192

EFFICIENCIES

Stage Efficiencies

Stages	Overall	Methane	n-Pentane	i-Pentane	n-Butane
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	i-Butane	Propane	Ethane	H2O
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000

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

Stage Efficiencies

Stages	Overall	i-Butane	Propane	Ethane	H2O
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]60*	NBP[0]77*	NBP[0]93*	NBP[0]113*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]135*	NBP[0]151*	NBP[0]170*	NBP[0]189*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]208*	NBP[0]227*	NBP[0]246*	NBP[0]264*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000

* Specified by user.

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

Stage Efficiencies

Stages	Overall	NBP[0]208*	NBP[0]227*	NBP[0]246*	NBP[0]264*
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]283*	NBP[0]302*	NBP[0]321*	NBP[0]340*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]359*	NBP[0]378*	NBP[0]397*	NBP[0]416*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]443*	NBP[0]484*	NBP[0]521*	NBP[0]562*
1__TS-1	1.000	1.000	1.000	1.000	1.000
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	0.5000
12__TS-1	1.000	1.000	1.000	1.000	1.000
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	0.3000

Stages	Overall	NBP[0]600*	NBP[0]641*	NBP[0]737*	
1__TS-1	1.000	1.000	1.000	1.000	

* Specified by user.

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

Stage Efficiencies

Stages	Overall	NBP[0]600*	NBP[0]641*	NBP[0]737*	
2__TS-1	0.5000 *	0.5000	0.5000	0.5000	
3__TS-1	0.5000 *	0.5000	0.5000	0.5000	
4__TS-1	0.5000 *	0.5000	0.5000	0.5000	
5__TS-1	0.5000 *	0.5000	0.5000	0.5000	
6__TS-1	0.5000 *	0.5000	0.5000	0.5000	
7__TS-1	0.5000 *	0.5000	0.5000	0.5000	
8__TS-1	0.5000 *	0.5000	0.5000	0.5000	
9__TS-1	0.5000 *	0.5000	0.5000	0.5000	
10__TS-1	0.5000 *	0.5000	0.5000	0.5000	
11__TS-1	0.5000 *	0.5000	0.5000	0.5000	
12__TS-1	1.000	1.000	1.000	1.000	
13__TS-1	0.3000 *	0.3000	0.3000	0.3000	
14__TS-1	0.3000 *	0.3000	0.3000	0.3000	

RATING

Tray Sections

Tray Section		TS-1		
Tray Diameter	(ft)	4.921		
Weir Height	(ft)	0.1640 *		
Weir Length	(ft)	3.937 *		
Tray Space	(ft)	1.640		
Tray Volume	(ft3)	31.20		
Disable Heat Loss Calculations		No		
Heat Model		None		
Rating Calculations		No		
Tray Hold Up	(ft3)	3.120		

CONDITIONS

Name		um column steam @Main	Vacuum Feed @Main	vacuum residue @Main	column overhead @Main
Vapour		1.0000	0.7545 *	0.0000	1.0000
Temperature	(F)	500.0000 *	760.0000 *	668.1414	149.9982
Pressure	(psia)	164.6960 *	1.3147	1.1989	0.9668
Molar Flow	(lbmole/hr)	1110.1798	1675.9674	494.2037	1241.1979
Mass Flow	(kg/h)	9071.9405 *	302078.3263	132506.8896	10813.1471
Std Ideal Liq Vol Flow	(kbpd)	1.3722	46.9450	19.6607	1.6567
Molar Enthalpy	(Btu/lbmole)	-1.007e+005	-1.928e+005	-3.539e+005	-1.036e+005
Molar Entropy	(Btu/lbmole-F)	41.32	337.0	454.9	48.17
Heat Flow	(Btu/hr)	-1.1177e+08	-3.2310e+08	-1.7490e+08	-1.2857e+08
Name		LVGO @Main	HVGO @Main	slop wax @Main	gePA_Q-Cooler_1 @Main
Vapour		0.0000	0.0000	0.0000	---
Temperature	(F)	373.2302	562.4118	684.3738	---
Pressure	(psia)	1.0204	1.0918	1.1453	---
Molar Flow	(lbmole/hr)	250.7361	769.4745	30.5350	---
Mass Flow	(kg/h)	29673.1112	131678.1547	6478.9642	---
Std Ideal Liq Vol Flow	(kbpd)	5.0000	20.9999	1.0000	---
Molar Enthalpy	(Btu/lbmole)	-2.024e+005	-2.487e+005	-2.714e+005	---
Molar Entropy	(Btu/lbmole-F)	119.7	240.6	347.2	---
Heat Flow	(Btu/hr)	-5.0739e+07	-1.9139e+08	-8.2874e+06	4.2000e+07 *

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

PROPERTIES

Name		uum column steam @M	Vacuum Feed @Main	vacuum residue @Main	um column overhead @	LVGO @Main
Molecular Weight		18.02	397.4	591.1	19.21	260.9
Molar Density	(lbmole/ft3)	1.660e-002	1.339e-004	8.394e-002	1.479e-004	0.1835
Mass Density	(lb/ft3)	0.2991	5.319e-002	49.62	2.840e-003	47.87
Act. Volume Flow	(barrel/day)	2.858e+005	5.352e+007	2.517e+004	3.588e+007	5842
Mass Enthalpy	(Btu/lb)	-5589	-485.2	-598.7	-5393	-775.6
Mass Entropy	(Btu/lb-F)	2.294	0.8482	0.7696	2.508	0.4588
Heat Capacity	(Btu/lbmole-F)	8.879	273.7	407.7	8.540	152.1
Mass Heat Capacity	(Btu/lb-F)	0.4928	0.6888	0.6897	0.4446	0.5831
LHV Molar Basis (Std)	(Btu/lbmole)	0.0000	---	---	---	---
HHV Molar Basis (Std)	(Btu/lbmole)	1.763e+004	---	---	---	---
HHV Mass Basis (Std)	(Btu/lb)	978.7	---	---	---	---
CO2 Loading		---	---	---	---	---
CO2 Apparent Mole Conc.	(lbmole/ft3)	---	---	---	---	---
CO2 Apparent Wt. Conc.	(lbmol/lb)	---	---	---	---	---
LHV Mass Basis (Std)	(Btu/lb)	---	---	---	---	---
Phase Fraction [Vol. Basis]		1.000	0.6510	0.0000	1.000	0.0000
Phase Fraction [Mass Basis]		1.000	0.6339	0.0000	1.000	0.0000
Phase Fraction [Act. Vol. Basis]		1.000	0.9996	0.0000	1.000	0.0000
Mass Exergy	(Btu/lb)	393.1	180.6	114.8	44.05	31.75
Partial Pressure of CO2	(psia)	0.0000	0.0000	0.0000	0.0000	0.0000
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000	0.0000
Act. Gas Flow	(ACFM)	1114	2.086e+005	---	1.399e+005	---
Avg. Liq. Density	(lbmole/ft3)	3.458	0.1526	0.1074	3.203	0.2144
Specific Heat	(Btu/lbmole-F)	8.879	273.7	407.7	8.540	152.1
Std. Gas Flow	(MMSCFD)	10.09	15.23	4.492	11.28	2.279
Std. Ideal Liq. Mass Density	(lb/ft3)	62.30	60.64	63.51	61.51	55.93
Act. Liq. Flow	(USGPM)	---	639.1	734.0	---	170.4
Z Factor		0.9632	---	1.180e-003	---	6.222e-004
Watson K		---	11.44	11.50	11.50	11.49
User Property		---	---	---	---	---
Partial Pressure of H2S	(psia)	0.0000	0.0000	0.0000	0.0000	0.0000
Cp/(Cp - R)		1.288	1.007	1.005	1.303	1.013
Cp/Cv		1.348	1.006	1.210	1.304	1.128
Heat of Vap.	(Btu/lbmole)	1.564e+004	2.446e+005	2.040e+005	1.932e+004	4.783e+004
Kinematic Viscosity	(cSt)	3.854	---	1.632	185.1	0.5908
Liq. Mass Density (Std. Cond)	(lb/ft3)	63.35	59.51	62.76	63.31	55.24
Liq. Vol. Flow (Std. Cond)	(barrel/day)	1349	4.784e+004	1.990e+004	1610	5062
Liquid Fraction		0.0000	0.2455	1.000	0.0000	1.000
Molar Volume	(ft3/lbmole)	60.23	7470	11.91	6762	5.450
Mass Heat of Vap.	(Btu/lb)	868.2	615.5	345.0	1006	183.3
Phase Fraction [Molar Basis]		1.0000	0.7545	0.0000	1.0000	0.0000
Surface Tension	(dyne/cm)	---	16.51	19.41	---	19.45
Thermal Conductivity	(Btu/hr-ft-F)	2.338e-002	---	5.432e-002	1.206e-002	6.072e-002
Viscosity	(cP)	1.847e-002	---	1.297	8.421e-003	0.4530
Cv (Semi-Ideal)	(Btu/lbmole-F)	6.893	271.7	405.7	6.554	150.2
Mass Cv (Semi-Ideal)	(Btu/lb-F)	0.3826	0.6838	0.6864	0.3412	0.5755
Cv	(Btu/lbmole-F)	6.589	272.2	336.9	6.549	134.8
Mass Cv	(Btu/lb-F)	0.3657	0.6849	0.5700	0.3410	0.5169
Cv (Ent. Method)	(Btu/lbmole-F)	---	---	368.3	---	139.3
Mass Cv (Ent. Method)	(Btu/lb-F)	---	---	0.6230	---	0.5340
Cp/Cv (Ent. Method)		---	---	1.107	---	1.092

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

PROPERTIES

Name	uum column steam @M	Vacuum Feed @Main	vacuum residue @Main	um column overhead @I	LVGO @Main
Reid VP at 37.8 C (psia)	0.9380	3.127e-004	5.162e-007	2.833e-002	2.084e-004
True VP at 37.8 C (psia)	0.9380	0.9382	0.3216	0.9673	0.1919
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	1349	4.763e+004	1.990e+004	1610	5062
Viscosity Index	---	-19.10	9.262	-8.206	-5.145
Name	HVGO @Main	slop wax @Main			
Molecular Weight	377.3	467.8			
Molar Density (lbmole/ft3)	0.1241	9.876e-002			
Mass Density (lb/ft3)	46.82	46.20			
Act. Volume Flow (barrel/day)	2.651e+004	1322			
Mass Enthalpy (Btu/lb)	-659.3	-580.2			
Mass Entropy (Btu/lb-F)	0.6377	0.7422			
Heat Capacity (Btu/lbmole-F)	249.0	329.0			
Mass Heat Capacity (Btu/lb-F)	0.6601	0.7033			
LHV Molar Basis (Std) (Btu/lbmole)	---	---			
HHV Molar Basis (Std) (Btu/lbmole)	---	---			
HHV Mass Basis (Std) (Btu/lb)	---	---			
CO2 Loading	---	---			
CO2 Apparent Mole Conc. (lbmole/ft3)	---	---			
CO2 Apparent Wt. Conc. (lbmol/lb)	---	---			
LHV Mass Basis (Std) (Btu/lb)	---	---			
Phase Fraction [Vol. Basis]	0.0000	0.0000			
Phase Fraction [Mass Basis]	0.0000	0.0000			
Phase Fraction [Act. Vol. Basis]	0.0000	0.0000			
Mass Exergy (Btu/lb)	80.42	121.6			
Partial Pressure of CO2 (psia)	0.0000	0.0000			
Cost Based on Flow (Cost/s)	0.0000	0.0000			
Act. Gas Flow (ACFM)	---	---			
Avg. Liq. Density (lbmole/ft3)	0.1566	0.1305			
Specific Heat (Btu/lbmole-F)	249.0	329.0			
Std. Gas Flow (MMSCFD)	6.995	0.2776			
Std. Ideal Liq. Mass Density (lb/ft3)	59.09	61.06			
Act. Liq. Flow (USGPM)	773.1	38.55			
Z Factor	8.022e-004	9.446e-004			
Watson K	11.49	11.50			
User Property	---	---			
Partial Pressure of H2S (psia)	0.0000	0.0000			
Cp/(Cp - R)	1.008	1.006			
Cp/Cv	1.138	1.161			
Heat of Vap. (Btu/lbmole)	6.296e+004	7.923e+004			
Kinematic Viscosity (cSt)	0.2372	0.2852			
Liq. Mass Density (Std. Cond) (lb/ft3)	58.38	60.34			
Liq. Vol. Flow (Std. Cond) (barrel/day)	2.126e+004	1012			
Liquid Fraction	1.000	1.000			
Molar Volume (ft3/lbmole)	8.059	10.13			
Mass Heat of Vap. (Btu/lb)	166.9	169.4			
Phase Fraction [Molar Basis]	0.0000	0.0000			
Surface Tension (dyne/cm)	17.01	15.89			
Thermal Conductivity (Btu/hr-ft-F)	5.558e-002	5.185e-002			
Viscosity (cP)	0.1779	0.2111			
Cv (Semi-Ideal) (Btu/lbmole-F)	247.1	327.0			
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.6549	0.6991			

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

PROPERTIES

Name		HVGO @Main	slop wax @Main			
Cv	(Btu/lbmole-F)	218.9	283.3			
Mass Cv	(Btu/lb-F)	0.5801	0.6055			
Cv (Ent. Method)	(Btu/lbmole-F)	228.8	298.3			
Mass Cv (Ent. Method)	(Btu/lb-F)	0.6066	0.6377			
Cp/Cv (Ent. Method)		1.088	1.103			
Reid VP at 37.8 C	(psia)	3.920e-006	2.410e-006			
True VP at 37.8 C	(psia)	0.1206	0.1370			
Liq. Vol. Flow - Sum(Std. Cond)	(barrel/day)	2.126e+004	1012			
Viscosity Index		-37.13	-25.42			

SUMMARY

Flow Basis:	Liquid Volume	The composition option is selected

Feed Composition

	Vacuum Feed	vacuum column steam	
Flow Rate (m3/h)	46.9450	1.3722	
	---	---	
Methane	0.0000	0.0000	
n-Pentane	0.0000	0.0000	
i-Pentane	0.0000	0.0000	
n-Butane	0.0000	0.0000	
i-Butane	0.0000	0.0000	
Propane	0.0000	0.0000	
Ethane	0.0000	0.0000	
H2O	0.0032	1.0000	
NBP[0]60*	0.0000	0.0000	
NBP[0]77*	0.0000	0.0000	
NBP[0]93*	0.0000	0.0000	
NBP[0]113*	0.0000	0.0000	
NBP[0]135*	0.0000	0.0000	
NBP[0]151*	0.0000	0.0000	
NBP[0]170*	0.0001	0.0000	
NBP[0]189*	0.0002	0.0000	
NBP[0]208*	0.0004	0.0000	
NBP[0]227*	0.0011	0.0000	
NBP[0]246*	0.0026	0.0000	
NBP[0]264*	0.0050	0.0000	
NBP[0]283*	0.0080	0.0000	
NBP[0]302*	0.0123	0.0000	
NBP[0]321*	0.0169	0.0000	
NBP[0]340*	0.0215	0.0000	
NBP[0]359*	0.0279	0.0000	
NBP[0]378*	0.0357	0.0000	
NBP[0]397*	0.0492	0.0000	
NBP[0]416*	0.0654	0.0000	
NBP[0]443*	0.1410	0.0000	
NBP[0]484*	0.1297	0.0000	
NBP[0]521*	0.1167	0.0000	
NBP[0]562*	0.0932	0.0000	
NBP[0]600*	0.0692	0.0000	
NBP[0]641*	0.0765	0.0000	
NBP[0]737*	0.1243	0.0000	

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

SUMMARY

Flow Basis: Liquid Volume | The composition option is selected

Feed Flows

	Vacuum Feed	vacuum column steam	
Flow Rate (m3/h)	46.9450	1.3722	
	---	---	
Methane (m3/h)	0.0000	0.0000	
n-Pentane (m3/h)	0.0000	0.0000	
i-Pentane (m3/h)	0.0000	0.0000	
n-Butane (m3/h)	0.0000	0.0000	
i-Butane (m3/h)	0.0000	0.0000	
Propane (m3/h)	0.0000	0.0000	
Ethane (m3/h)	0.0000	0.0000	
H2O (m3/h)	0.1504	1.3722	
NBP[0]60* (m3/h)	0.0000	0.0000	
NBP[0]77* (m3/h)	0.0000	0.0000	
NBP[0]93* (m3/h)	0.0001	0.0000	
NBP[0]113* (m3/h)	0.0002	0.0000	
NBP[0]135* (m3/h)	0.0006	0.0000	
NBP[0]151* (m3/h)	0.0015	0.0000	
NBP[0]170* (m3/h)	0.0033	0.0000	
NBP[0]189* (m3/h)	0.0078	0.0000	
NBP[0]208* (m3/h)	0.0185	0.0000	
NBP[0]227* (m3/h)	0.0499	0.0000	
NBP[0]246* (m3/h)	0.1199	0.0000	
NBP[0]264* (m3/h)	0.2324	0.0000	
NBP[0]283* (m3/h)	0.3761	0.0000	
NBP[0]302* (m3/h)	0.5770	0.0000	
NBP[0]321* (m3/h)	0.7946	0.0000	
NBP[0]340* (m3/h)	1.0114	0.0000	
NBP[0]359* (m3/h)	1.3089	0.0000	
NBP[0]378* (m3/h)	1.6779	0.0000	
NBP[0]397* (m3/h)	2.3102	0.0000	
NBP[0]416* (m3/h)	3.0718	0.0000	
NBP[0]443* (m3/h)	6.6173	0.0000	
NBP[0]484* (m3/h)	6.0888	0.0000	
NBP[0]521* (m3/h)	5.4786	0.0000	
NBP[0]562* (m3/h)	4.3760	0.0000	
NBP[0]600* (m3/h)	3.2474	0.0000	
NBP[0]641* (m3/h)	3.5900	0.0000	
NBP[0]737* (m3/h)	5.8344	0.0000	

Products

Flow Basis: Liquid Volume | The composition option is selected

Product Compositions

	vacuum column overhead	LVGO	HVGO
Flow Rate (m3/h)	1.6567	5.0000	20.9999
	---	---	---
Methane	0.0000	0.0000	0.0000
n-Pentane	0.0000	0.0000	0.0000
i-Pentane	0.0000	0.0000	0.0000
n-Butane	0.0000	0.0000	0.0000
i-Butane	0.0000	0.0000	0.0000
Propane	0.0000	0.0000	0.0000

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

SUMMARY

	vacuum column overhead	LVGO	HVGO
Ethane	0.0000	0.0000	0.0000
H2O	0.9189	0.0000	0.0000
NBP[0]60*	0.0000	0.0000	0.0000
NBP[0]77*	0.0000	0.0000	0.0000
NBP[0]93*	0.0001	0.0000	0.0000
NBP[0]113*	0.0001	0.0000	0.0000
NBP[0]135*	0.0003	0.0000	0.0000
NBP[0]151*	0.0009	0.0000	0.0000
NBP[0]170*	0.0020	0.0000	0.0000
NBP[0]189*	0.0046	0.0000	0.0000
NBP[0]208*	0.0106	0.0002	0.0000
NBP[0]227*	0.0226	0.0025	0.0000
NBP[0]246*	0.0230	0.0162	0.0000
NBP[0]264*	0.0106	0.0425	0.0001
NBP[0]283*	0.0038	0.0729	0.0003
NBP[0]302*	0.0016	0.1122	0.0006
NBP[0]321*	0.0007	0.1522	0.0015
NBP[0]340*	0.0002	0.1826	0.0046
NBP[0]359*	0.0001	0.1865	0.0179
NBP[0]378*	0.0000	0.1185	0.0516
NBP[0]397*	0.0000	0.0577	0.0961
NBP[0]416*	0.0000	0.0309	0.1385
NBP[0]443*	0.0000	0.0210	0.3077
NBP[0]484*	0.0000	0.0037	0.2717
NBP[0]521*	0.0000	0.0003	0.1082
NBP[0]562*	0.0000	0.0000	0.0011
NBP[0]600*	0.0000	0.0000	0.0000
NBP[0]641*	0.0000	0.0000	0.0000
NBP[0]737*	0.0000	0.0000	0.0000

	slop wax	vacuum residue	
Flow Rate (m3/h)	1.0000	19.6607	
	---	---	
Methane	0.0000	0.0000	
n-Pentane	0.0000	0.0000	
i-Pentane	0.0000	0.0000	
n-Butane	0.0000	0.0000	
i-Butane	0.0000	0.0000	
Propane	0.0000	0.0000	
Ethane	0.0000	0.0000	
H2O	0.0000	0.0000	
NBP[0]60*	0.0000	0.0000	
NBP[0]77*	0.0000	0.0000	
NBP[0]93*	0.0000	0.0000	
NBP[0]113*	0.0000	0.0000	
NBP[0]135*	0.0000	0.0000	
NBP[0]151*	0.0000	0.0000	
NBP[0]170*	0.0000	0.0000	
NBP[0]189*	0.0000	0.0000	
NBP[0]208*	0.0000	0.0000	
NBP[0]227*	0.0000	0.0000	
NBP[0]246*	0.0000	0.0000	

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

SUMMARY

	slop wax	vacuum residue	
NBP[0]264*	0.0000	0.0000	
NBP[0]283*	0.0001	0.0000	
NBP[0]302*	0.0002	0.0000	
NBP[0]321*	0.0003	0.0000	
NBP[0]340*	0.0005	0.0000	
NBP[0]359*	0.0010	0.0000	
NBP[0]378*	0.0018	0.0000	
NBP[0]397*	0.0038	0.0000	
NBP[0]416*	0.0077	0.0001	
NBP[0]443*	0.0349	0.0008	
NBP[0]484*	0.1405	0.0114	
NBP[0]521*	0.5669	0.1341	
NBP[0]562*	0.1916	0.2117	
NBP[0]600*	0.0393	0.1632	
NBP[0]641*	0.0109	0.1820	
NBP[0]737*	0.0005	0.2967	

Flow Basis: Liquid Volume The composition option is selected

Product Flows

	vacuum column overhead	LVGO	HVGO
Flow Rate (m3/h)	1.6567 *	5.0000 *	20.9999 *
	---	---	---
Methane (m3/h)	0.0000 *	0.0000 *	0.0000 *
n-Pentane (m3/h)	0.0000 *	0.0000 *	0.0000 *
i-Pentane (m3/h)	0.0000 *	0.0000 *	0.0000 *
n-Butane (m3/h)	0.0000 *	0.0000 *	0.0000 *
i-Butane (m3/h)	0.0000 *	0.0000 *	0.0000 *
Propane (m3/h)	0.0000 *	0.0000 *	0.0000 *
Ethane (m3/h)	0.0000 *	0.0000 *	0.0000 *
H2O (m3/h)	1.5223 *	0.0001 *	0.0001 *
NBP[0]60* (m3/h)	0.0000 *	0.0000 *	0.0000 *
NBP[0]77* (m3/h)	0.0000 *	0.0000 *	0.0000 *
NBP[0]93* (m3/h)	0.0001 *	0.0000 *	0.0000 *
NBP[0]113* (m3/h)	0.0002 *	0.0000 *	0.0000 *
NBP[0]135* (m3/h)	0.0006 *	0.0000 *	0.0000 *
NBP[0]151* (m3/h)	0.0015 *	0.0000 *	0.0000 *
NBP[0]170* (m3/h)	0.0033 *	0.0000 *	0.0000 *
NBP[0]189* (m3/h)	0.0076 *	0.0001 *	0.0000 *
NBP[0]208* (m3/h)	0.0175 *	0.0009 *	0.0001 *
NBP[0]227* (m3/h)	0.0374 *	0.0123 *	0.0002 *
NBP[0]246* (m3/h)	0.0381 *	0.0810 *	0.0008 *
NBP[0]264* (m3/h)	0.0175 *	0.2127 *	0.0022 *
NBP[0]283* (m3/h)	0.0064 *	0.3644 *	0.0053 *
NBP[0]302* (m3/h)	0.0026 *	0.5612 *	0.0130 *
NBP[0]321* (m3/h)	0.0011 *	0.7610 *	0.0322 *
NBP[0]340* (m3/h)	0.0004 *	0.9132 *	0.0972 *
NBP[0]359* (m3/h)	0.0001 *	0.9324 *	0.3754 *
NBP[0]378* (m3/h)	0.0000 *	0.5925 *	1.0835 *
NBP[0]397* (m3/h)	0.0000 *	0.2883 *	2.0179 *
NBP[0]416* (m3/h)	0.0000 *	0.1545 *	2.9084 *
NBP[0]443* (m3/h)	0.0000 *	0.1052 *	6.4620 *
NBP[0]484* (m3/h)	0.0000 *	0.0186 *	5.7059 *

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

SUMMARY

	vacuum column overhead	LVGO	HVGO
NBP[0]521* (m3/h)	0.0000 *	0.0016 *	2.2731 *
NBP[0]562* (m3/h)	0.0000 *	0.0000 *	0.0224 *
NBP[0]600* (m3/h)	0.0000 *	0.0000 *	0.0001 *
NBP[0]641* (m3/h)	0.0000 *	0.0000 *	0.0000 *
NBP[0]737* (m3/h)	0.0000 *	0.0000 *	0.0000 *

	slop wax	vacuum residue	
Flow Rate (m3/h)	1.0000 *	19.6607 *	
	---	---	
Methane (m3/h)	0.0000 *	0.0000 *	
n-Pentane (m3/h)	0.0000 *	0.0000 *	
i-Pentane (m3/h)	0.0000 *	0.0000 *	
n-Butane (m3/h)	0.0000 *	0.0000 *	
i-Butane (m3/h)	0.0000 *	0.0000 *	
Propane (m3/h)	0.0000 *	0.0000 *	
Ethane (m3/h)	0.0000 *	0.0000 *	
H2O (m3/h)	0.0000 *	0.0002 *	
NBP[0]60* (m3/h)	0.0000 *	0.0000 *	
NBP[0]77* (m3/h)	0.0000 *	0.0000 *	
NBP[0]93* (m3/h)	0.0000 *	0.0000 *	
NBP[0]113* (m3/h)	0.0000 *	0.0000 *	
NBP[0]135* (m3/h)	0.0000 *	0.0000 *	
NBP[0]151* (m3/h)	0.0000 *	0.0000 *	
NBP[0]170* (m3/h)	0.0000 *	0.0000 *	
NBP[0]189* (m3/h)	0.0000 *	0.0000 *	
NBP[0]208* (m3/h)	0.0000 *	0.0000 *	
NBP[0]227* (m3/h)	0.0000 *	0.0000 *	
NBP[0]246* (m3/h)	0.0000 *	0.0000 *	
NBP[0]264* (m3/h)	0.0000 *	0.0000 *	
NBP[0]283* (m3/h)	0.0001 *	0.0000 *	
NBP[0]302* (m3/h)	0.0002 *	0.0000 *	
NBP[0]321* (m3/h)	0.0003 *	0.0000 *	
NBP[0]340* (m3/h)	0.0005 *	0.0000 *	
NBP[0]359* (m3/h)	0.0010 *	0.0000 *	
NBP[0]378* (m3/h)	0.0018 *	0.0001 *	
NBP[0]397* (m3/h)	0.0038 *	0.0003 *	
NBP[0]416* (m3/h)	0.0077 *	0.0012 *	
NBP[0]443* (m3/h)	0.0349 *	0.0153 *	
NBP[0]484* (m3/h)	0.1405 *	0.2238 *	
NBP[0]521* (m3/h)	0.5669 *	2.6370 *	
NBP[0]562* (m3/h)	0.1916 *	4.1620 *	
NBP[0]600* (m3/h)	0.0393 *	3.2080 *	
NBP[0]641* (m3/h)	0.0109 *	3.5791 *	
NBP[0]737* (m3/h)	0.0005 *	5.8338 *	

Flow Basis: Liquid Volume The composition option is selected

Product Recoveries

	vacuum column overhead	LVGO	HVGO
Flow Rate (m3/h)	1.6567	5.0000	20.9999
	---	---	---
Methane (%)	99.9939	0.0018	0.0041
n-Pentane (%)	99.9449	0.0279	0.0263
i-Pentane (%)	99.9494	0.0250	0.0248

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

SUMMARY

	vacuum column overhead	LVGO	HVGO
n-Butane (%)	99.9666	0.0152	0.0175
i-Butane (%)	99.9708	0.0129	0.0157
Propane (%)	99.9803	0.0080	0.0113
Ethane (%)	99.9882	0.0042	0.0073
H2O (%)	99.9790	0.0034	0.0065
NBP[0]60* (%)	99.9147	0.0458	0.0383
NBP[0]77* (%)	99.8877	0.0630	0.0478
NBP[0]93* (%)	99.8511	0.0872	0.0599
NBP[0]113* (%)	99.7890	0.1304	0.0785
NBP[0]135* (%)	99.6864	0.2054	0.1056
NBP[0]151* (%)	99.5591	0.3033	0.1344
NBP[0]170* (%)	99.2850	0.5312	0.1798
NBP[0]189* (%)	98.4838	1.2676	0.2435
NBP[0]208* (%)	94.6061	5.0529	0.3343
NBP[0]227* (%)	74.9127	24.6113	0.4674
NBP[0]246* (%)	31.7870	67.5432	0.6585
NBP[0]264* (%)	7.5221	91.5157	0.9473
NBP[0]283* (%)	1.6904	96.8766	1.4132
NBP[0]302* (%)	0.4504	97.2759	2.2467
NBP[0]321* (%)	0.1372	95.7734	4.0523
NBP[0]340* (%)	0.0407	90.2965	9.6104
NBP[0]359* (%)	0.0099	71.2360	28.6784
NBP[0]378* (%)	0.0015	35.3097	64.5762
NBP[0]397* (%)	0.0001	12.4773	87.3469
NBP[0]416* (%)	0.0000	5.0300	94.6811
NBP[0]443* (%)	0.0000	1.5891	97.6531
NBP[0]484* (%)	0.0000	0.3063	93.7118
NBP[0]521* (%)	0.0000	0.0289	41.4905
NBP[0]562* (%)	0.0000	0.0001	0.5115
NBP[0]600* (%)	0.0000	0.0000	0.0033
NBP[0]641* (%)	0.0000	0.0000	0.0000
NBP[0]737* (%)	0.0000	0.0000	0.0000

	slop wax	vacuum residue	
Flow Rate (m3/h)	1.0000	19.6607	
	---	---	
Methane (%)	0.0002	0.0000	
n-Pentane (%)	0.0009	0.0000	
i-Pentane (%)	0.0008	0.0000	
n-Butane (%)	0.0006	0.0000	
i-Butane (%)	0.0006	0.0000	
Propane (%)	0.0004	0.0000	
Ethane (%)	0.0003	0.0000	
H2O (%)	0.0003	0.0109	
NBP[0]60* (%)	0.0012	0.0000	
NBP[0]77* (%)	0.0014	0.0000	
NBP[0]93* (%)	0.0017	0.0000	
NBP[0]113* (%)	0.0021	0.0000	
NBP[0]135* (%)	0.0027	0.0000	
NBP[0]151* (%)	0.0032	0.0000	
NBP[0]170* (%)	0.0041	0.0000	
NBP[0]189* (%)	0.0052	0.0000	

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

SUMMARY

	slop wax	vacuum residue	
NBP[0]208* (%)	0.0066	0.0000	
NBP[0]227* (%)	0.0086	0.0000	
NBP[0]246* (%)	0.0112	0.0000	
NBP[0]264* (%)	0.0148	0.0000	
NBP[0]283* (%)	0.0198	0.0000	
NBP[0]302* (%)	0.0269	0.0001	
NBP[0]321* (%)	0.0370	0.0002	
NBP[0]340* (%)	0.0519	0.0005	
NBP[0]359* (%)	0.0743	0.0013	
NBP[0]378* (%)	0.1087	0.0038	
NBP[0]397* (%)	0.1637	0.0120	
NBP[0]416* (%)	0.2511	0.0378	
NBP[0]443* (%)	0.5270	0.2308	
NBP[0]484* (%)	2.3068	3.6752	
NBP[0]521* (%)	10.3479	48.1326	
NBP[0]562* (%)	4.3785	95.1100	
NBP[0]600* (%)	1.2093	98.7874	
NBP[0]641* (%)	0.3036	99.6964	
NBP[0]737* (%)	0.0094	99.9906	

COLUMN PROFILES

Reflux Ratio:	1.343	Reboil Ratio:	---	The Flows Option is Selected	Flow Basis: Liquid Volume

Column Profiles Flows

	Temp (F)	Pres (psia)	Net Liq (kbpd)	Net Vap (kbpd)	Net Feed (kbpd)	Net Draws (kbpd)	Duty (Btu/hr)
1__TS-1	150.0	1	30.81	---	22.30	1.657	-5.37e+007 *
2__TS-1	239.0	1	36.68	10.16	---	---	---
3__TS-1	321.4	1	44.89	16.03	---	---	---
4__TS-1	373.2	1	18.35	24.25	---	27.30	---
5__TS-1	462.3	1	83.92	25.00	50.00	---	-6.49e+007 *
6__TS-1	499.1	1	94.36	40.58	---	---	---
7__TS-1	526.7	1	98.87	51.02	---	---	---
8__TS-1	562.4	1	26.80	55.53	---	71.00	---
9__TS-1	631.9	1	27.03	54.45	---	---	---
10__TS-1	665.4	1	26.49	54.69	---	---	---
11__TS-1	684.4	1	18.71	54.15	---	1.000	---
12__TS-1	717.2	1	26.63	47.37	46.94	---	---
13__TS-1	693.0	1	22.39	8.343	---	---	---
14__TS-1	668.1	1	---	4.099	1.372	19.66	---

Column Profiles Energy

	Temperature (F)	Liq Enthalpy (Btu/lbmole)	Vap Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
1__TS-1	150.0	-2.135e+005	-1.036e+005	---
2__TS-1	239.0	-1.972e+005	-1.142e+005	---
3__TS-1	321.4	-1.911e+005	-1.158e+005	---
4__TS-1	373.2	-2.024e+005	-1.229e+005	---
5__TS-1	462.3	-2.431e+005	-1.293e+005	---
6__TS-1	499.1	-2.386e+005	-1.407e+005	---
7__TS-1	526.7	-2.399e+005	-1.470e+005	---
8__TS-1	562.4	-2.487e+005	-1.504e+005	---
9__TS-1	631.9	-2.634e+005	-1.529e+005	---

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

COLUMN PROFILES

	Temperature (F)	Liq Enthalpy (Btu/lbmole)	Vap Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
10__TS-1	665.4	-2.665e+005	-1.542e+005	---
11__TS-1	684.4	-2.714e+005	-1.532e+005	---
12__TS-1	717.2	-3.093e+005	-1.450e+005	---
13__TS-1	693.0	-3.336e+005	-1.182e+005	---
14__TS-1	668.1	-3.539e+005	-1.083e+005	---

FEEDS / PRODUCTS

Flow Basis: Liquid Volume

	Stream	Type	Duty (kJ/h)	State	Flows (m3/h)	Enthalpy (kJ/kgmole)	Temp (C)
1__TS-1	<PA_1>	Energy	-5.4e+007 *		---	---	---
	PA_1_Return	Feed	---	Liquid	22.3 *	-2.5e+005 *	-10.47
	vacuum column overh	Draw	---	Vapour	1.66 *	-1.0e+005 *	150.0
2__TS-1							
3__TS-1							
4__TS-1	LVGO	Draw	---	Liquid	5.00 *	-2.0e+005 *	373.2
	PA_1_Draw	Draw	---	Liquid	22.3 *	-2.0e+005 *	373.2
5__TS-1	<HVGO pump around	Energy	-6.5e+007 *		---	---	---
	HVGO pump around_	Feed	---	Liquid	50.0 *	-2.8e+005 *	412.4
6__TS-1							
7__TS-1							
8__TS-1	HVGO	Draw	---	Liquid	21.0 *	-2.5e+005 *	562.4
	HVGO pump around_	Draw	---	Liquid	50.0 *	-2.5e+005 *	562.4
9__TS-1							
10__TS-1							
11__TS-1	slop wax	Draw	---	Liquid	1.00 *	-2.7e+005 *	684.3
12__TS-1	Vacuum Feed	Feed	---	Mixed	46.9 *	-1.9e+005 *	759.9
13__TS-1							
14__TS-1	vacuum column stean	Feed	---	Vapour	1.37 *	-1.0e+005 *	500.0
	vacuum residue	Draw	---	Liquid	19.7 *	-3.5e+005 *	668.1

SETUP

Column Flowsheet Topology

Total Theor. Stages:	14 *	Total Tray-Sections:	1 *	Condenser + Reboiler:	0 *	Pump Arounds:	2 *
Side Strippers:	0 *	Side Rectifiers:	0 *	Vapour Bypasses:	0 *		

Sub-Flowsheet

Internal Feed Stream	External Feed Stream	Transfer Basis
vacuum column steam	vacuum column steam @Main	P-H Flash
Vacuum Feed	Vacuum Feed @Main	P-H Flash
Internal Prod Stream	**External Prod Stream**	**Transfer Basis**
vacuum column overhead	vacuum column overhead @Main	P-H Flash
vacuum residue	vacuum residue @Main	P-H Flash
TopStagePA_Q-Cooler_1	TopStagePA_Q-Cooler_1 @Main	None Req'd
LVGO	LVGO @Main	P-H Flash
HVGO	HVGO @Main	P-H Flash
slop wax	slop wax @Main	P-H Flash
HVGO pump around_Q-Cooler		None Req'd
LVGO_Draw		P-H Flash
PA_1_Q-Cooler		None Req'd

* Specified by user.

Appendix (C)

Aspen Hysys: Cost Report

Aspen Hysys: Cost Report

PROJSUM.ICS (Project Summary)		
ITEM	**UNITS**	**VALUE**
***		
PROJECT INFORMATION		
Project Name		6.0 ATMOSPHERIC CRUDE OIL
Project Description		
Analysis Date and Time		Sat Oct 17 00:04:44 2015
Simulator Type		Aspen HYSYS
Simulator Version		
Simulator Report File		C:\USERS
Simulator Report Date		Thursday, June 11, 2015
Economic Analysis Type		IPE
Version		21.0.0
Project Directory		C:\USERS
Scenario Name		Scenario1
Scenario Description		
***		
CAPITAL COST EVALUATION BASIS		
Date		17-oct-15
Country		US
Units of Measure		I-P
Currency (Cost) Symbol		U.S. DOLLAR
Currency Conversion Rate	USD/U.S. DOLLAR	1
System Cost Base Date		1Q 12
Project Type		Grass roots/Clear field
Design code		ASME
Prepared By		
Plant Location		North America
Capacity		1.#INF
Time Difference Between System Cost Base Date and Start Date for Engineering	Days	-820
User Currency Name		DOLLARS
User Currency Description		U.S. DOLLARS
User Currency Symbol		USD
***		
TIME PERIOD		
Period Description		Year
Operating Hours per Period	Hours/period	8000
Number of Weeks per Period	Weeks/period	52
Number of Periods for Analysis	Period	20

**		
SCHEDULE		
Start Date for Engineering		01-janv-10
Duration of EPC Phase	Weeks	69
Length of Start-up Period	Weeks	20
Duration of Construction Phase	Weeks	56
Completion Date for Construction		Thursday, May 05, 2011
**		
CAPITAL COSTS PARAMETERS		
Working Capital Percentage	Percent/period	5
**		
OPERATING COSTS PARAMETERS		
Operating Supplies (lump-sum)	Cost/period	0
Laboratory Charges (lump-sum)	Cost/period	0
User Entered Operating Charges (as percentage)	Percent/period	25
Operating Charges (Percent of Operating Labor Costs)	Percent/period	25
Plant Overhead (Percent of Operating Labor and Maintenance Costs)	Percent/period	50
G and A Expenses (Percent of Subtotal Operating Costs)	Percent/period	8
**		
GENERAL INVESTMENT PARAMETERS		
Tax Rate	Percent/period	40
Interest Rate	Percent/period	20
Economic Life of Project	Period	10
Salvage Value (Fraction of Initial Capital Cost)	Percent	20
Depreciation Method		Straight Line
**		
ESCALATION		
Project Capital Escalation	Percent/period	5
Products Escalation	Percent/period	5
Raw Material Escalation	Percent/period	3,5

Operating and Maintenance Labor Escalation	Percent/period	3
Utilities Escalation	Percent/period	3
**		
PROJECT RESULTS SUMMARY		
Total Project Capital Cost	Cost	7,17E+07
Total Raw Materials Cost	Cost/period	0
Total Products Sales	Cost/period	0
Total Operating Labor and Maintenance Cost	Cost/period	2,66E+06
Total Utilities Cost	Cost/period	2,55E+07
Total Operating Cost	Cost/period	3,20E+07
Operating Labor Cost	Cost/period	600000
Maintenance Cost	Cost/period	2,06E+06
Operating Charges	Cost/period	150000
Plant Overhead	Cost/period	1,33E+06
Subtotal Operating Cost	Cost/period	2,96E+07
G and A Cost		2,37E+06
**		
PROJECT CAPITAL SUMMARY		Total Cost
Purchased Equipment	Cost	2,65E+07
Equipment Setting	Cost	681922
Piping	Cost	1,07E+07
Civil	Cost	3,49E+06
Steel	Cost	400759
Instrumentation	Cost	1,57E+06
Electrical	Cost	540796
Insulation	Cost	2,41E+06
Paint	Cost	128280
Other	Cost	1,81E+07
Subcontracts	Cost	0
G and A Overheads	Cost	1,81E+06
Contract Fee	Cost	2,17E+06
Escalation	Cost	0
Contingencies	Cost	1,23E+07
Total Project Cost	Cost	8,08E+07
Adjusted Total Project Cost	Cost	7,17E+07
**		
ENGINEERING SUMMARY		Cost
Basic Engineering		915800
Detail Engineering		1,86E+06

Material Procurement		578500
Home Office		885500
Total Design, Eng, Procurement Cost		4,24E+06
***		
RAW MATERIALS COSTS AND PRODUCTS SALES		
Raw Materials Cost per Hour	Cost/Hour	0
Total Raw Materials Cost	Cost/Period	0
Products Sales per Hour	Cost/Hour	0
Total Products Sales	Cost/Period	0
Main Product Name		
Main Product Rate		0
Main Product Unit Cost	1.#INF	0
Main Product Production Basis		
Main Product Rate per Period	1.#INF	0
Main Product Sales	USD/Year	0
By-product Sales	USD/Year	0
***		
OPERATING LABOR AND MAINTENANCE COSTS		
Operating Labor		
Operators per Shift		2
Unit Cost	Cost/Operator/H	20
Total Operating Labor Cost	Cost/period	320000
Maintenance		
Cost/8000 Hours		2,06E+06
Total Maintenance Cost	Cost/period	2,06E+06
Supervision		
Supervisors per Shift		1
Unit Cost	Cost/Supervisor/H	35
Total Supervision Cost	Cost/period	280000
***		
UTILITIES COSTS		
Electricity		
Rate	KW	127,665
Unit Cost	Cost/KWH	0,0775
Total Electricity Cost	Cost/period	79152,3
Potable Water		

Rate		
Unit Cost	Cost/MMGAL	0
Total Potable Water Cost	Cost/period	0
Fuel		
Rate		
Unit Cost	Cost/MMBTU	7,85
Total Fuel Cost	Cost/period	0
Instrument Air		
Rate		
Unit Cost	Cost/KCF	0
Total Instrument Air Cost	Cost/period	0
Subtotal Cost	Cost/period	79152,3
Process Utilities Subtotal Cost	Cost/period	2,54E+07

Aspen Hysys: Cost Report

PROJECT CAPITAL SUMMARY	**Design, Eng, Procurement**	**Construction Material**	**Construction Manhours**	**Construction Manpower**	**Construction In directs**
Purchased Equipment		2,65E+07			
Equipment Setting			22279	681922	
Piping		7,31E+06	111734	3,38E+06	
Civil		2,06E+06	58179	1,42E+06	
Steel		345164	1992	55594,5	
Instrumentation		1,37E+06	6620	200994	
Electrical		460543	2781	80253,3	
Insulation		1,23E+06	52531	1,19E+06	
Paint		41818,3	3866	86461,8	
Other	4,24E+06	4,03E+06			9,81E+06
Subcontracts					
G and A Overheads	0	1,30E+06		212811	294321
Contract Fee	220527	892101		445698	616406
Escalation	0	0		0	0
Contingencies	803057	8,19E+06		1,40E+06	1,93E+06
Adjusted Total Project Cost					
********************************					
ENGINEERING SUMMARY	Manhours				
Basic Engineering	8252				
Detail Engineering	17484				
Material Procurement					
Home Office	8726				