DESIGN AND FABRICATION OF CRUDE DISTILLATION UNIT COMPONENTS FOR A MINI PETROLEUM REFINERY.

By

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DECLARATION

I declare that the work in the thesis entitled 'DESIGN AND FABRICATION OF CRUDE DISTILLATION UNIT COMPONENTS FOR A MINI PETROLEUM REFINERY.' has been performed by me at the Department of Chemical Engineering under the supervision of Dr. I A Mohammed-Dabo and Dr. A Hamza. The information derived from the literature has been duly acknowledged in the text and a list of references provided. No part of this thesis was previously presented for another degree or diploma at any university.

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Name of Student	Signature	Date

CERTIFICATION

This thesis entitled 'DESIGN AND FABRICATION OF CRUDE DISTILLATION UNIT COMPONENTS FOR A MINI PETROLEUM REFINERY.' by Oche Adoyi meets the regulations governing the award of the degree of Master of Science in Chemical Engineering of Ahmadu Bello University, and is approved for its contribution to knowledge and literary presentation.

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ABSTRACT

Nigeria is faced with high import of petroleum products, incessant shortage of the products and poor petroleum refining status. There is therefore the need to develop local technology in petroleum refining as a sustainable solution to these problems. This work contributes its own quota to this worthy course by designing and fabricating crude distillation unit components for a mini petroleum refinery to process 1 bbl/day of Escravos crude. The unit consists of atmospheric distillation unit and vacuum distillation unit. Components designed and fabricated are: an Atmospheric distillation column, a Vacuum distillation column, an ADU furnace, a VDU furnace, a Condenser, a Reflux drum, 3 Strippers, an AGO-Crude Heat exchanger, a Top Pump-around heat exchanger and a Bottom Pump-around heat exchanger.

Design was done using both manual calculations and ASPEN Plus software. The ASPEN Plus simulation gave: Internal diameter of ADU, VDU and Stripper columns as 7.5cm, 9.9cm and 2cm respectively. Required number of trays for ADU, VDU and Stripper columns was found to be 20, 9 and 4 trays respectively. ADU furnace was determined to be a double section box furnace with 12 tubes in convection section and 5 tubes in radiation section while VDU furnace had 4 tubes in radiation section only. All tubes are 0.4m long with diameter of 5mm. ADU top product condenser required was a shell and tube heat exchanger with 3 tubes each of which had a length of 0.5m and diameter 5mm while crude-AGO, top pump around and bottom pump around heat exchangers were double pipe with tube lengths 0.3m, 0.2m and 0.2m respectively. Reflux drum required is vertical separator of height x diameter; 32cm x 5cm respectively. Working drawing of each of the designed components was prepared for ease of fabrication.

All components were successfully fabricated and ready for installation.

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CHAPTER ONE

1.0 **INTRODUCTION**

1.1 Preamble

Petroleum being a mixture of hydrocarbons has a boiling range of -160^{0}C (methane) to 1000^{0}C or more (pitch) i.e. to say a mixture of gas, liquid and solid, requires an effective and economic distillation to process into a number of cuts of small boiling range. These cuts are later processed and tailored to suit the requirement of consumers.

Petroleum refining are processes that convert petroleum into refinery products such as petrol (gasoline), kerosene, diesel and lube oil. The processes are classified into three namely: distillation (separation based differences in product boiling range), cracking (breaking less valuable heavy hydrocarbon to more valuable light hydrocarbon) and treatment (removal of fuel contaminants). Yield of refinery products from a refinery depends on the effectiveness of this processes and the petroleum feed stock. Light crude yields high amount of light product such as petrol. Approximately 65% of the crude oil produced in Nigeria is light (35^{0}API or higher) and sweet (low sulphur content).

The basis of refinery distillation design rests completely on TBP tests. Distillation of crude mainly takes place in two stages. First stage distillation is carried out at atmospheric pressure; hence the name 'Atmospheric Distillation Unit' (ADU) is conferred on it. The undistillated portion of crude, called reduced crude is further distilled under reduced pressure in a second unit known as 'Vacuum Distillation Unit' (VDU) (Rao, 1990).

The world crude oil refining capacity is approximately 82 million barrels per calendar day (b/cd) equivalent to about 4,200 million tones (Mt) per annum. This capacity is currently provided by a total of about 720 refineries. Many of these refineries are old

and out-dated, some produce only specialist products such as lubricating oils and asphalt, and many are small and uneconomic.

Some predictions of growth suggest that the world will be consuming 90 or more million barrels of oil per day by 2015, compared to 75 million now. With existing refinery capacity reaching its practical limits, this requires the equivalent of up to 100 new world-scale refineries.

Nigeria has four (4) Crude oil refineries located in Port Harcourt, Warri and Kaduna, with a total nameplate refining capacity of 445,000BPD. Port Harcourt has two refineries, one with a capacity of 60,000BPD and the other 150,000BPD, Warri Refinery – 125,000BPD and Kaduna Refinery – 110,000BPD.

Nigeria which is the largest crude oil exporter in Africa remains the largest importer of petroleum products in the continent. (http://tribune.come.ng/index.php/editorial/46129-reawakening-the-ghost-of-fuel-subsidy-removal). The country continues to experience perpetual shortages of products due in part to poor configurations and inefficient operations of the refineries, resulting in frequent breakdowns occasioned by poor or lack of Turn Around Maintenance (TAM).

In the first quarter of year 2005, the Nigerian market consumed over 700, 000BPD of refined products out of which over half were imported. Furthermore, there is a substantial West African export market with over 1, 000, 000BPD market demand being serviced by less than 600, 000BPD in nameplate refining capacity from the various refineries in the region (Nkaginieme, 2005).

This offers considerable opportunity for private and efficiently run refineries. The Nigerian Government has already issued about 15 ‘Approval to Construct’ licences for refineries. Most of these licenses are above 50 000BPD capacity. Implementation will

prove to be extremely difficult due to the large financial outlay involved- estimated at between $0.5 billion and $1.0 billion. The licenced small refineries are however more likely to come on-stream sooner due to the manageable financial requirements (Nkaginieme, 2005). This prompted this research that seeks to address local development of mini petroleum refining technology. The petroleum feed stock to be used was Escravos crude from Delta state, Nigeria. 100% of the components of the mini refinery were fabricated locally. The Mini Petroleum Refinery is to consist of two units: atmospheric distillation unit (ADU) and vacuum distillation unit (VDU) and its refining capacity is 159 litre per day (1 BPD) of crude oil feed.

1.2 Research Problems

The work has the following research problems:

I. Literatures on mini refinery design based on Nigerian crudes are not readily available.

II. No Nigerian tertiary institution has a functional mini-refinery for purpose of teaching, research and development.

III. All the four refineries in Nigeria were designed and constructed by foreign companies. This leads to the spending of huge amount of money to the original designers for the Turn-Around- Maintenance or repairs of some faulty units.

1.3 Research Scopes

The research scopes for the work are:

I. Design components of atmospheric distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

II. Design components of vacuum distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

III. Fabricate components of atmospheric distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

IV. Fabricate components of vacuum distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

1.4 Research Aim and Objectives:

1.4.1 Aim

The main aim of the work is to design and construct components for a Mini Petroleum Refinery (Crude Distillation Unit).

1.4.2 Objectives

This work has the following objectives:

I. To design components of atmospheric distillation unit for a Mini refinery of capacity159 litres per day using Escravos crude as feedstock.

II. To design components of vacuum distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

III. To fabricate components of atmospheric distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

IV. To fabricate components of vacuum distillation unit for a Mini refinery of capacity 159 litres per day using Escravos crude as feedstock.

1.5 Justification

The work the following justification:

I. Looking at Nigeria's history, we have had issues of petroleum product supply ranging from fuel shortage to fuel price increment to fuel subsidy removal and then subsidy probe. The only sure and sustainable way to remedies this situation is local development of technology in Mini petroleum refinery.

II. Academic mini refinery would allow local development of the technology and its scale up to commercial capacity, thereby building local technological capacity (capacity in terms of building, operation and maintenance of refinery).

III. It enhances the development and consolidation of indigenous human capital in the country's most strategic sector because the academic refineries will serve as avenues of practical experience to students in relevant disciplines.

IV. The work also provides a platform for better utilization of our material resources in process fabrications.

V. It will motivate the higher education sector, which now finds a practical role in the development process, to seek more avenues to contribute to national development.

VI. It will increase the supply of petroleum products in the long term.

VII. It will reduce the amount of foreign exchange spent on importation of petroleum products.

VIII. Scale up (expansion) of the academic refineries will set the stage for Nigeria to earn value-added foreign currency through the export of petroleum products to other countries.

IX. Work will increase local content in the oil and gas sector both human and material.

X. There has not been any report of an academic mini petroleum refinery anywhere in Nigeria.

1.6 Expected Contributions to Knowledge

The expected contributions to knowledge for the work are:

I. This research would greatly help Nigeria develop local capacity in the design and fabrication of process plants through pilot plant and scale up in critical areas of technology (such as petroleum refining) for national self-sufficiency, self-reliance and sustainable growth.

II. The models developed would help not only in scale up but optimization of the entire petroleum refining process.

CHAPTER TWO

2.0 LITERATURE SURVEY

2.1 Petroleum

Petroleum is a substance occurring naturally in the earth in solid, liquid, or gaseous state and composed mainly of mixtures of chemical compounds of carbon and hydrogen, with or without other nonmetallic elements such as sulfur, oxygen, and nitrogen. In some cases, especially in the measurement of oil and gas, petroleum refers only to oil- a liquid hydrocarbon- and does not include natural gas or gas liquids such as propane and butane. Bulk of petroleum is made up of saturated compounds like paraffins, naphthenes and unsaturated cyclic compounds mainly aromatics (Rao, 1990).

2.2 Petroleum Refining

The petroleum refining industry converts crude oil into more than 2500 refined products, including liquefied petroleum gas, gasoline, kerosene, aviation fuel, diesel fuel, fuel oils, lubricating oils, and feedstocks for the petrochemical industry. Petroleum refinery activities start with receipt of crude for storage at the refinery, include all petroleum handling and refining operations, and they terminate with storage preparatory to shipping the refined products from the refinery (Burklin, 1977).

The petroleum refining industry employs a wide variety of processes. A refinery's processing flow scheme is largely determined by the composition of the crude oil feedstock and the chosen slate of petroleum products. The example refinery flow scheme presented in Figure 2.1 shows the general processing arrangement.

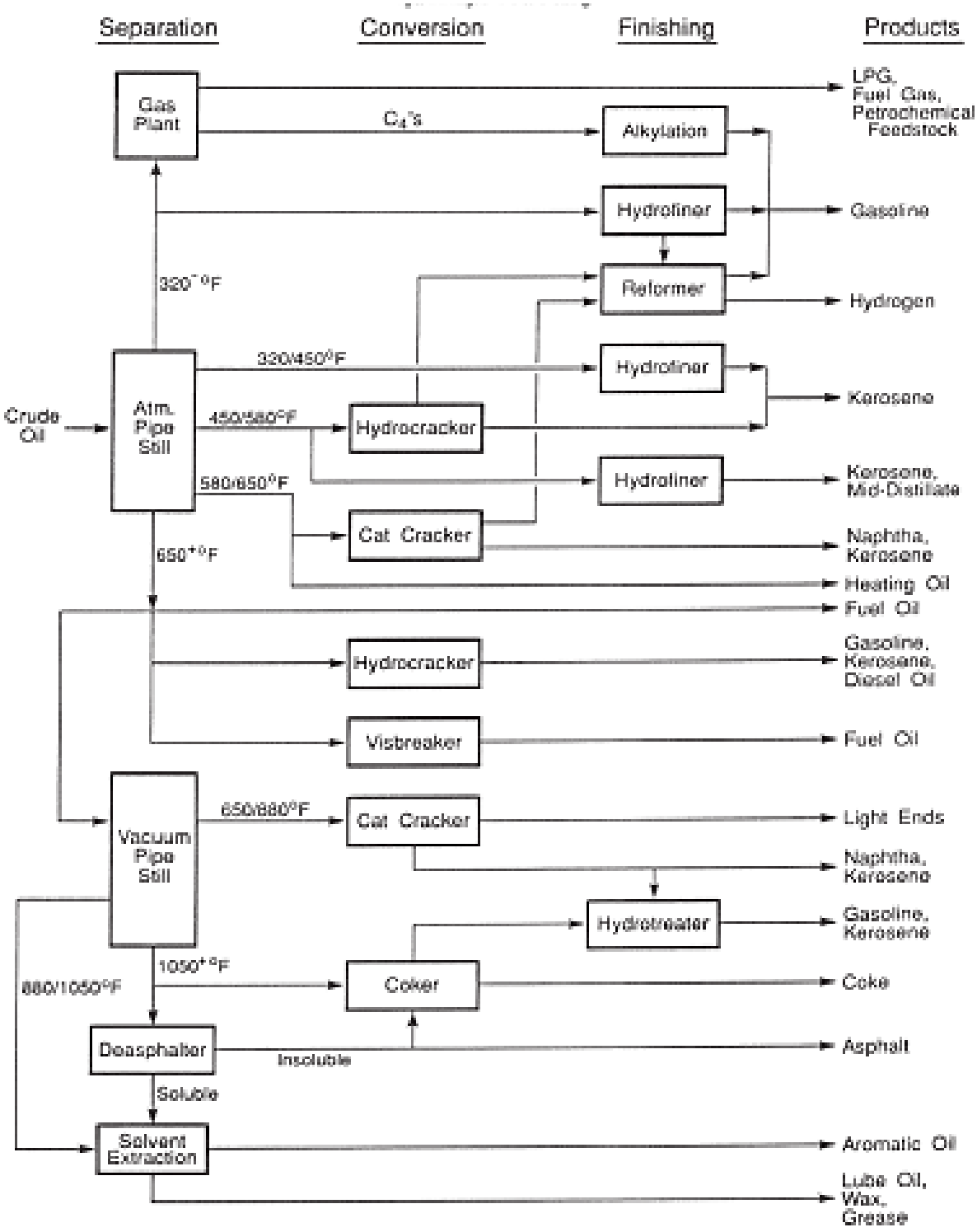


Figure 2.1: Typical Refinery Process Flow showing Separation, Conversion and Finishing Sections with End products (Eman, 2013).

Listed below are 5 categories of general refinery processes and associated operations:

2.2.1 Separation Processes –

The first phase in petroleum refining operations is the separation of crude oil into its major constituents using 3 petroleum separation processes: atmospheric distillation, vacuum distillation, and light ends recovery (gas processing). Crude oil consists of a mixture of hydrocarbon compounds including paraffinic, naphthenic, and aromatic hydrocarbons with small amounts of impurities including sulfur, nitrogen, oxygen, and metals. Refinery separation processes separate these crude oil constituents into common boiling-point fractions (Burklin, 1977).

2.2.2 Conversion Processes –

To meet the demands for high-octane gasoline, jet fuel, and diesel fuel, components such as residual oils, fuel oils, and light ends are converted to gasolines and other light fractions. Cracking, coking, and visbreaking processes are used to break large petroleum molecules into smaller ones.Polymerization and alkylation processes are used to combine small petroleum molecules into larger ones. Isomerization and reforming processes are applied to rearrange the structure of petroleum molecules to produce higher-value molecules of a similar molecular size (Burklin, 1977).

2.2.3 Treating Processes –

Petroleum treating processes stabilize and upgrade petroleum products by separating them from less desirable products and by removing objectionable elements. Undesirable elements such as sulfur, nitrogen, and oxygen are removed by hydrodesulfurization, hydrotreating, chemical sweetening, and acid gas removal. Treating processes, employed primarily for the separation of petroleum products, include such processes as deasphalting. Desalting is used to remove salt, minerals, grit, and water from crude oil

feedstocks before refining. Asphalt blowing is used for polymerizing and stabilizing asphalt to improve its weathering characteristics (Burklin, 1977).

2.2.4 Feedstock and Product Handling –

The refinery feedstock and product handling operations consist of unloading, storage, blending, and loading activities (Burklin, 1977).

2.2.5 Auxiliary Facilities –

A wide assortment of processes and equipment not directly involved in the refining of crude oil is used in functions vital to the operation of the refinery. Examples are boilers, waste water treatment facilities, hydrogen plants, cooling towers, and sulfur recovery units. Products from auxiliary facilities (clean water, steam, and process heat) are required by most process units throughout the refinery (Burklin, 1977).

2.3 Overview of African Refining

There are a total of 42 refineries in Africa, with a total name-plate capacity of 3,217,600 barrels per day (bpd). The major refining countries are Egypt with 9 refineries (774,900 bpd); Algeria with 5 refineries (303,700 bpd); Libya with 5 refineries (380,000 bpd); South Africa with 4 refineries (545,000); and Nigeria with 3 refineries (445,000 bpd).

More than half (i.e. 51.7%) of Africa's refining capacity is in North Africa. According to McKinsey's database, most of these are State-Owned (59%). 29% are based on joint ownership with government and 12% are joint-venture arrangements between International Oil Companies (IOC's). (www.vanguardngr.com/2014/03/crude-refining-africa-way-forward/ *accessed 2014).*

These national refineries operate at different levels of efficiency. Considering data of average national capacity utilizations from 2006-2009, the efficiency levels for these countries are: Egypt (81%), Algeria (94%), Libya (87%), South Africa (85%), and

Nigeria (18%). (www.vanguardngr.com/2014/03/crude-refining-africa-way-forward/ *accessed 2014).*

In the past 20 years, only 3 Greenfield refineries have been constructed in Africa. These were built in Adrar (Algeria) and Khartoum (Sudan) with China National Petroleum Company (CNPC) partnering with the governments, with capacities of 13,000 bpd and 100,000 bpd respectively.

The third one was built in Alexandria (Egypt) by Egypt General Petroleum Corporation, Egypt's National Oil Company (NOC) with a capacity of 100,000 bpd. Planned new builds were constructed by Petrochina at Ndjamena (Chad) and Zinder (Niger) with same 20,000 bpd capacity. The third is being constructed by Sonangol, Angola's NOC at Lobito (Angola) with a capacity of 200,000 bpd. From the foregoing, refining in Africa is led by NOC's, and new investments are dominated by the Chinese National Petroleum Companies. (*www.vanguardngr.com/2014/03/crude-refining-africa-way-forward/ accessed 2014)*. Table 2.1 shows the major refineries in Africa.

Table 2.1 Major Refineries in Africa

South Africa has 4 refineries and 3 synfuels plants.
1) Caltex has a 5.5 million mt/yr (110,000 b/d) refinery in Cape Town.
2) Shell and BP have joint ownership of the 8.2 million mt per annum (165,000 b/d) Saprefrefinery in Durban.
3) Engen has the 6 million mt/yr (125,000 b/d) Enref refinery in Durban.
4) Sasol and Total have joint ownership of the 4.2 million mt/yr (85,000 b/d) Natref refinery in Sasolburg.
Nigeria has three refineries, all owned by the Nigerian National Petroleum Company, NNPC. The Nigerian government has announced its intention to sell 51% of each of the refineries in 2004.
1) Kaduna refinery in northern Nigeria was built in 1980 with a capacity of 5.5 million mt/yr (110,000 b/d). A Lube Base oil plant was added in 1982, and an

LAB plant in 1987.
2) Warri refinery in the south central region was built in 1978 with a capacity of 6.2 million mt/yr (125,000 b/d). A Carbon Black plant and a Polypropylene plant were added in 1986
3) Port Harcourt refinery in the southeast is made up of two refineries, built in 1965 and 1989. In 1993 they were merged into one, with a total capacity of 10.500 million mt/yr (210,000 b/d).
Egypt has 9 refineries mostly concentrated in the northeast (Cairo, Alexandria, Suez). Egyptian General Petroleum Corp (EGPC) operates all but one of the refineries. The exception is the MIDOR Refinery in Alexandria.
1) El Mex refinery in Alexandria is operated by the Alexandria Petroleum Company. It has a capacity of 100,000 b/d, and 22,500 b/d of vacuum distillation capacity. In addition it has a Lube Baseoil manufacturing plant and a Bitumen unit.
2) Cairo Petroleum Refining Company in Mostorod, near Cairo. This refinery has a capacity of 145,000b/d.
3) The El-Nasr Petroleum Company near Suez has a capacity of 99,300 b/d. It has a 35,000 b/d Hydrocracker and a Bitumen unit.
4) The Amiriyah Petroleum Refining Company in Alexandria has a capacity of 78,000 b/d, and a 15,000 b/d vacuum distillation unit. It has a 9,000 b/d Alkylation unit, and a 2,000 b/d lube baseoil manufacturing unit.
5) The Suez Petroleum Processing Company near Suez has a capacity of 66,400 b/d, and a 9,500 b/d vacuum distillation unit. It has a 16,400 b/d Delayed Coker, and a 1,000 b/d Lube Baseoil unit.
6) The Asyut Petroleum Refining Company near the center of Egypt has a capacity of 47,000 b/d. This simple refinery has a small Naphtha Reformer, and is designed to supply product to the central and southern regions.
7) The Tanta refinery near Port Said is operated by the Cairo Petroleum Refining Company. It has a capacity of 35,000 b/d. Other than a small Hydrotreating unit it has no upgrading capacity.
8) The El-Nasr Petroleum Company operates the small Wadi Feran refinery on the Red Sea in the Gulf of Suez. It has capacity of 7,000 b/d, and was designed to service operations related to the Suez Canal.
9) The Middle East Oil Refinery (MIDOR) was completed in 2002 in the Amiriyah Free Zone, Alexandria. It has a capacity of 100,000 b/d, and has a 35,000 b/d Hydrocracker, a 22,800 b/d Coker, and a 10,700 b/d Isomerisation unit. This is

the only privately owned refinery in Egypt. It was originally a joint Egyptian/Israeli venture, but the Israeli shareholders sold out to the Egyptian National Bank in 2001.
Algeria is another major refining centre with 4 refineries.
1) Algiers refinery, built in 1964, has a capacity of 60,000 b/d. It was built by CFP/Total to supply the main market. It has no cracking capacity, and no special units.
2) Arzew refinery, built in 1973, has a capacity of 60,000 b/d. It was built by Sonatrach as an export refinery. It has no cracking facilities, but has Lube Baseoil (1984 built), and Bitumen units.
3) Skikda refinery, built in 1980 (expanded in 1993), has a capacity of 300,000 b/d. It was built by Sonatrach as an export refinery and petrochemical complex. It has no cracking facilities, but has an Aromatics unit (BTX) and a Bitumen plant.
4) Hassi Messaoud, a small refinery in southern Algeria, built in 1960, and expanded in 1979, has a capacity of 30,000 b/d.

Source: (*www.vanguardngr.com/2014/03/crude-refining-africa-way-forward/ accessed 2014).*

2.4 Overview of Nigerian Refining

Nigeria's state-held refineries (Port Harcourt I and II, Warri, and Kaduna) have a combined nameplate capacity of 438,750 bbl/d, but problems including sabotage, fire, poor management and lack of regular maintenance contribute to a low current capacity of around 214,000 bbl/day. Plans for several small, independently-owned refineries are also being developed, with the Nigerian government planning for three new refineries to come onstream by 2008. This has presently not being achieved. The $1.5-billion Tonwei refinery in Bayelsa State appears set to be the first private refinery in Nigeria, with a planned initial capacity of 100,000 bbl/d (Nkaginieme, 2005). The only Refinery built in Nigeria in the past 15 years is the 1,000 BOPD Ogbele Diesel Palnt owned by Niger Delta Petroleum Company located in Bayelsa State.

The U.S. Export-Import Bank has agreed to provide a loan guarantee for $10 million of the $29.8 million total cost of the 12,000-bbl/d Amake Modular Refinery in Akwa Ibom state, located in Eket. Production at Phase 1 of the refinery is expected to begin by mid-2005. Phase 2 of the Amake refinery will allow for another 6,000-bbl/d distillation unit, expected to be operational in mid-2006.

The NNPC's four oil refineries, its numerous petrochemicals plants, and its Pipelines and Products Marketing Company (PPMC) are due to be sold in Nigeria's continuing effort to privatize state entities. Although the Nigerian government planned to offer 51 percent stakes in each of its four refineries, attempts to privatize the Port Harcourt refinery in 2004 elicited no bids from the oil majors. Because multinational oil companies have shown little interest in investing in refinery privatization, the Nigerian government recently opened negotiations with Libyan, Indian, and Chinese investors (Nkaginieme, 2005).

2.5 Important Refinery Products

2.5.1 Refinery off gas

In a refinery, gas is formed in cracking and reforming operations due to the thermal degradation of liquid hydrocarbons. During stabilization of wild gasolines or processed gasolines, the gases are vented. Thus the gas is mainly a mixture of saturates and unsaturates and quantity is also not assessable. This forms a major source of heat energy for refinery, as well as feedstock for petrochemicals. In fact, without any exclusion, all these gases can be utilized for petrochemical industries.

2.5.2 Liquefied petroleum gas (LPG)

The gas that is vented from refinery distillation units, is processed and conveniently stored as propane-butane blends after liquefaction. For domestic heating purposes, it is

supplied in small cylinders 15kg or 12kg. While for industrial tanker supplies are called in.

2.5.3 Gasoline

Naphtha or Gasoline is the next fraction to follow gas. The boiling point ranges from 37^0C to 180^0C. Gasoline is a finished product, while raw fraction is termed as naphtha or light boiling fraction. There are different types of gasolines produced by the refineries (approx.. 40 types) and almost 90% of the product is exclusively consumed by automobile industry and the rest by aviation industry. Most refineries produce two primary grades, regular and premium, both either leaded or unleaded.

2.5.4 Kerosene

Kerosene is the general name applied to the group of refined petroleum fractions employed as fuel and illuminant. All these fractions have approximate boiling range 150-250^0C. These are uniform close cut distillates, low in viscosity, with a good degree of refinement to be fairly stable, light in color and free from smoky ill smelling substances. It is highly desirable that sulphur be kept as low as possible. Kerosene is used as illuminating oil in domestic needs (Rao, 1990).

2.5.5 Aviation turbine fuel (ATF), jet fuels

Modern jet engines use fuel similar to kerosene. It is a most flexible fuel in its boiling range (up to 300^0C). All properties which are desirable for kerosene are considered to be sufficient. Pour point of this fuel is of extreme importance and should not be higher than -30^0C as international flights always visit lands of caprice climates.

2.5.6 Diesel fuels

Diesel oils are the fractions in the boiling range of 250-320^{0}C: and fall under gas oil fractions. These are basically divided into two classes as high speed and low speed diesels. Classification of diesel oils is done according to speed and loads of the engine.

2.5.7 Lube oils

The principal source of lubricating oil is the fraction that is left after lighter components, namely gasoline, kerosene, kerosene, diesel oil during crude distillation. Generally lubes have a boiling point above 350^{0}C and these are obtained as the main products from vacuum distillation units. Residuums, after precipitation of asphaltenes are known as bright stocks and form a good source for lube oils. Depending upon the services intended for, lube oils are classified. Nelson grouped the oils into seven categories as shown below:

1) Machine and Engine oils
2) Compounded oils
3) Turbine oils
4) Cold test oils
5) Transformer oils
6) Colour oils
7) Corrosive oils

2.5.8 Bitumen

Bitumen is the residual product obtained from crude distillation unit. It is essentially solid at room temperature and has got very high viscosity. Asphalt is usually a mixture of bitumen in oil, containing much mineral matter. Bitumen obtained from the distillation column is poor in qualities. This has necessitated air blowing of bitumen to

obtain suitable grades. Most of the bitumen is used in high way constructions, waterproofing and coating works (Rao, 1990).

2.6 Mini-Refinery

Mini-refinery are refineries that have feedstock refining capacity of less than 20, 000 bbl/day. The feedstock varies from a single or a blend of the following: light crude, heavy crude, long residue and condensates. The units for a particular mini-refinery depends on the feedstock, desired products, desired product quality and availability of utilities. The uses of mini-refinery are listed below:

I. For research and development.

II. Test run reconstruction of existing refinery.

III. Supply of refinery products to remote areas.

IV. They supplement product supply from larger commercial refineries. (www.scawa.com)

Some photographs of mini-refinery around the world as obtained from the internet are as presented in Figure 2.2 and Figure 2.3:



Figure 2.2: 150BPD Mini-refinery, Papua New Guinea. (www.minirefinery.com accessed 2013)



Figure 2.3: 150BPD Mini-refinery, Siberia Russia. (www.minirefinery.com accessed 2013)

2.7 Crude Distillation Unit

Crude Distillation Units are key process plants in a petroleum refinery as they produce intermediate streams that are used in downstream process units. Changes in these units have a great impact on product yield and quality and, therefore, it is recommended to operate these units at optimal conditions from technical and economical points of view; that means operating

conditions such as temperatures, pressures and flows of the units that maximize their economic performance (increasing product yield), subject to their real physical restrictions and their design capabilities (Lopez, 2009).

Crude distillation is energy intensive. It consumes fuel at the equivalent of 2% of the crude processed. The conventional design (Figure 2.4), consisting of a column with side strippers and pump-around circuits, appeared 70 years ago and is still the design used in the refining industry. Because energy efficiency is now desired, all of these designs merit reevaluations. Nevertheless, the conventional design is widespread and popular (Bagajewicz, 2001).

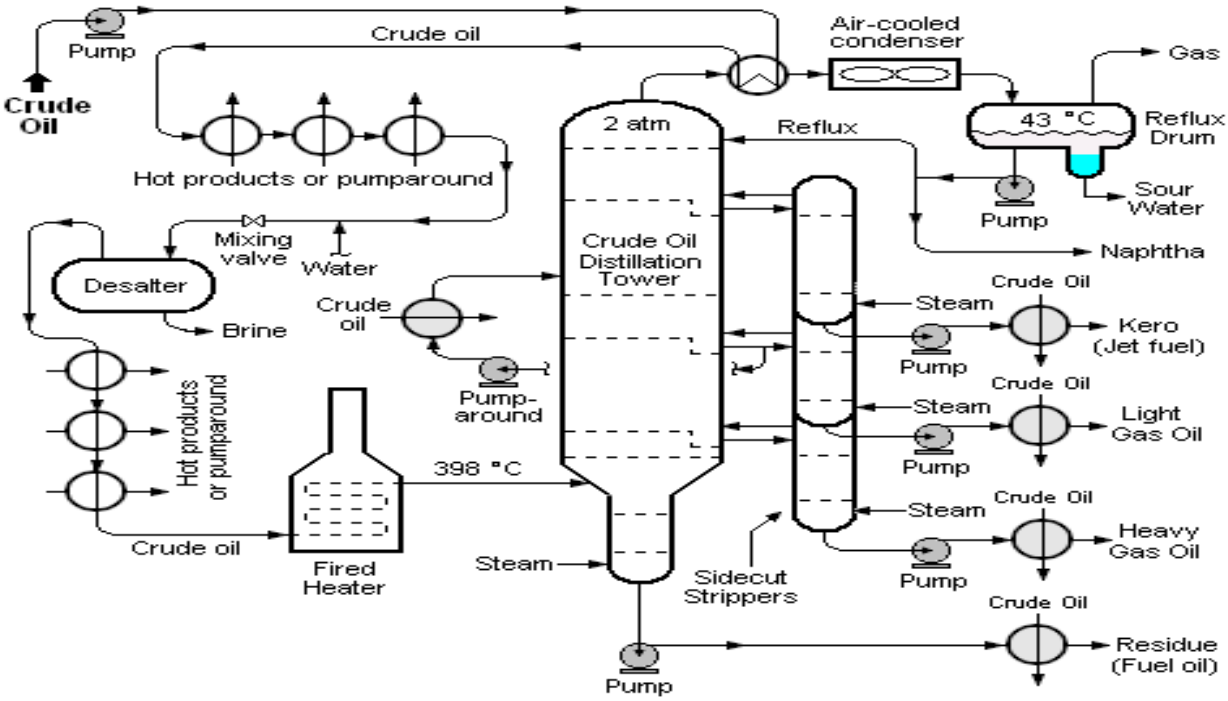


Figure 2.4: Crude distillation Unit (Massimiliano, 2011)

Crude is mixed with water and heated in a heat exchanger network before entering a desalter, where most of the water containing the salt is removed. The desalted crude enters another heat exchanger network and receives heat from hot streams. Both heat exchanger networks make use of the vapors of the main column condenser, the pump-around circuit streams, and the products that need to be cooled. The preheated crude then enters the furnace, where it is heated to about 340-370 °C. The partially vaporized crude is fed into the flash zone of the atmospheric column, where the vapor and liquid separate. The vapor includes all of the components that comprise the products, while the liquid is the residue with a small amount of relatively light components in the range of gas oil. These components are removed from the residue by steam stripping at the bottom of the column. In addition to the overhead condenser, there are several pump-around circuits along the column, where liquid streams are withdrawn, cooled, and sent

back to upper trays. Products are withdrawn in the liquid state from different trays and then stripped by steam in side strippers to remove light components (Bagajewicz, 2001).

Crude oil is a complex mixture. There exist about 1000 distinguishable components with boiling temperatures varying from room temperature to over 550°C. Crude distillation yields mixtures called naphtha, kerosene, diesel, and gas oil. These products are specified by ASTM D86 distillation temperatures. Compared to common distillation of discrete components, crude distillation has the following specific features:

(1) Large processing quantity: The charge rate is the largest among all petroleum or chemical processing units. The typical processing capability is around 15 000 m^3 per day (100 000 bbl/day). In such a large scale process, energy cost accounts for a larger part of manufacturing costs than in other processes.

(2) Large temperature variation throughout the column: The temperature difference between the top tray and the flash zone is about 250°C, which means significant heat degradation throughout the column.

(3) Absence of a reboiler: The main column functions as a rectifying section for products, while side columns act as stripping sections.

(4) Low separation sharpness: Product quality is specified by ASTM boiling points rather than component fractions as in the discrete component separation case. The former is a more relaxed requirement.

(5) Components in a lighter product can be found in any of heavier products: This is because all components constituting the light product have to travel through trays where heavier products are withdrawn (Bagajewicz, 2001).

The major objective in the design of crude distillation units is to find the most energy-efficient separation structure. Although some ideas exist for the design of energy-integrated distillation schemes, they are not directly applicable to crude fractionation for the following reasons:

(1) The number of components in the crude is too large to handle. Usually around 30-40 pseudo components are used, while available studies on sequencing seldom addressed systems containing over 5 components.

(2) Previous separation sequencing studies assumed that the products were pure; however, products in crude distillation are mixtures.

2.8 Parts of the Crude Distillation Unit

2.8.1 Desalting Unit

All crudes contain moisture and salts to varying degrees. Water is likely to occur in emulsion from when the crudes are naphthenic or sulfurous. No harm maybe expected to the distillation column due to the presence of moisture, as there is always stream in distillation. However, crude have to be dehydrated to remove the salts. Water being good solvent for these salts, the removal is very much effective in the form of brine. Of all the existing salts, chlorides of calcium and magnesium distinguish themselves in playing an invincible role in overhead corrosion to the equipment. Any crude that contain more than 5kg of total salts expressed in terms of sodium chloride per thousand barrels may be regarded as salty crude (Rao, 1990).

In general, dehydration of crude is practiced in two stages; first at the site of mine and later in the refinery. The following general methods are versatile for dehydration of crudes.

1. Chemical treatment
2. Gravity settling
3. Centrifugal separation
4. Electric desalter

At the oil field, salt is removed by settling or by adding chemicals or by combination of these two. Crudes possessing emulsifying characteristics are not responsive to the settling method. The settling techniques are also not effective and time consuming. Continuous operations are not possible with the result large amount of spaced and equipments are to be isolated for this purpose. Similarly centrifugal separation is also not economical due to the huge energy requirements and less quantity handled; all these have given the way to electric desalting (Rao, 1990).

Electric Desalting

Simultaneous desalting and dehydration is achieved in this unit with a spectacular removal of more than 90% salt in just less than half an hour. The principle in this separation is very simple; under a charged electric field the polar molecules orient. A potential of 20000 to 30000 volts is applied between electrodes through which crude is passed. Water present in the form of emulsion also coalesces and agglomerates into a stream entrapping all the salts in this process. Brine collects at the bottom of the desalter, while crude floats above and forms a separates stream (Rao, 1990).

Compactness, efficiency and ease of operation of these units are indubitable and induced every refinery to adopt. Temperature during electric desalting is maintained

around 90^0C and a pressure of 5 to 6 kg/cm^2 is also superimposed. Power consumption is also very small, frequently of the order0.01 kWhr per barrel.

2.8.2 Crude Preheating Unit

Before the introduction of pipe still heaters (1918) into refinery operations, the direct heating stills (convective type) were used for all heating operations. Nonuniform heating and excessive heating could not be abated satisfactorily in these stills; with these inescapable difficulties, the refinery could not be magnified or utilized for high capacities at short notice, thus one has to be content with small refinery sizes. The introduction of pipe still heaters, and other chemical engineering appliances into refinery, has tremendously increased the activities in the trade (Rao, 1990).

Pipe still heaters are different from old still heaters in one way; the radiant section is separated from convection section. Prudent planning of radiant and convection sections are primary considerations for a pipe still heater for efficacious operations (Rao, 1990).

2.8.3 Heat Exchangers and Steam Reboilers

Heat exchangers are also used to preheat the feedstock before it enters the furnace. These exchangers are bundles of tubes arranged within a shell so that a stream passes through the tubes in the opposite direction of a stream passing through the shell. Thus cold crude oil, by passing through a series of heat exchangers where hot products from the distillation tower are cooled, before entering the furnace and saving of heat in this manner, may be a major factor in the economical operation of refineries (Speight, 1998).

Steam reboilers may take the form of a steam coil at the bottom of the fractional distillation tower or in a separate vessel. In the latter case, the bottom product from the tower enters the reboiler where part is vaporized by heat from the steam coil. The hot vapour is directed back to the bottom of the tower and provides part of the heat needed to operate the tower. The non-volatile product leaves the reboiler and passes through a heat exchanger, where its heat is tranferred to the feed to the tower. Steam may also be inject ted into a fractional distillation tower, not only to provide heat but also to induce boiling to take place at lower temperatures. Reboilers generally increase the efficiency of fractionation (Speight, 1998).

2.9 Plant Design

The general term plant design includes all engineering aspects involved in the development of either a new, modified, or expanded industrial plant. In this development, the chemical engineer will be making economic evaluations of new processes, designing individual pieces of equipment for the proposed new venture, or developing a plant layout for coordination of the overall operation. Because of these many design duties, the chemical engineer is many times referred to here as a design engineer. In many instances, the term process engineering is used in connection with economic evaluation and general economic analyses of industrial processes, while process design refers to the actual design of the equipment and facilities necessary for carrying out the process (Timmerhaus, 1991).

Various types of computer programs and techniques are used to carry out the design of individual pieces of equipment or to develop the strategy for a full plant design. This application of computer usage in design is designated as *computer-aided design.*

The chemical engineer has many tools to choose from in the development of a profitable plant design. None, when properly utilized, will probably contribute as much to the optimization of the design as the use of high-speed computers. Many problems encountered in the process development and design can be solved rapidly with a higher degree of completeness with high-speed computers and at less cost than with ordinary hand or desk calculators. Generally overdesign and safety factors can be reduced with a substantial savings in capital investment.

At no time, however, should the engineer be led to believe that plants are designed around computers. They are used to determine design data and are used as models for optimization once a design is established. They are also used to maintain operating plants on the desired operating conditions. The latter function is a part of design and supplements and follows process design.

In the engineer's approach to any design problem, it is necessary to be prepared to make many assumptions. Sometimes these assumptions are made because no absolutely accurate values or methods of calculation are available. At other times, methods involving close approximations are used because exact treatments would require long and laborious calculations giving little gain in accuracy. The good chemical engineer recognizes the need for making certain assumptions but also knows that this type of approach introduces some uncertainties into the final results. Therefore, assumptions are made only when they are necessary and essentially correct (Timmerhaus, 1991).

2.10: Past Related Works

Table 2.3 show the past related works consulted, work done by authors and conclusions from the work.

Table 2.3: Past Related works

	Work Done	Conclusion
Akpa, 2013	Steady state models for a multi-component crude distillation column were developed from the MESH equations. The accuracy of the models was ascertained with plant data of the crude distillation unit of the Port-Harcourt Refinery.	The model results of the concentrations and temperatures for five components compared favourably with output values from the distillation unit with maximum deviations of 8.33% and 6.62% respectively. The developed model also accurately predicted the tray of withdrawal of the five components/fractions in the distillation column.
Ji, 2002	Rigorous targeting procedures were performed for three types of complete crude distillation plants.	It was found that the introduction of a vacuum tower changes the topologies for both the conventional design and the preflash design, thereby changing the heat distribution among the pump-around circuits.
Kamel, 2013	A retrofit design approach was developed for existing crude oil distillation systems with their associated heat exchanger networks. The approach is a rigorous simulation and optimisation-based model that considers simultaneously the existing distillation column with its full details and the associated HEN.	Several retrofit solutions was obtained, ranging from zero-modifications and simple additional exchanger areas to additional units or equipments. Simple optimisations of both the distillation operation and the HEN have lead to 17% savings in energy consumptions and emissions. The addition of a preflash to existing structure showed large energy savings of up to 32 % compared with base case and substantial utility cost savings per year of 1,065,678 $. The payback time for most retrofit solutions did not exceed a couple of months. Atmospheric CO2 emissions have been reduced significantly with retrofit by up to 33% with respect to emissions of 18,679 kg/h for the base case.
Lekan et al, 2013	This research work investigated the expert system design and control of crude oil distillation column (CODC) using artificial neural network model which was validated	Artificial neural network model is an effective tool for the design and control of crude oil distillation column.

	using experimental data obtained from functioning crude oil distillation column of Port-Harcourt Refinery, Nigeria.	
Mohammed, 2002.	Design was done for a 20, 000 bbl/day Medium Size Refinery II: Upgrading of a Hydro-skimming Plant.	Upgrading Medium size Refinery of 20, 000 bbl/day was found to be a viable project in Nigeria.
Manasseh, 2001.	Design of a Medium size Refinery to Process 20,000 barrels per stream day.	Medium size Refinery of capacity 20,000 bbl/day is an economically viable project in Nigeria.
Massimiliano et al, 2008	In this work we attention was focused to an industrial crude oil distillation unit, evaluating the possibility to modify the feed conditions by installing a preflash drum or a preflash plate column. Real data plant were collected to obtain a reliable simulation of the unit by means of the software package Aspen Plus 13.0.	The results obtained were compared with the plant data in terms of flow rate and product quality utilizing the ASTM D-86 curves and a good agreement was obtained.
Nuhu, 2013	The energy analysis was developed with Aspen Hysys software using design and operational data of crude distillation unit of N'djamena Refinery.	The ideal work, lost and shaft work were 2.40E+08, 4.29E+08 and 6.69E+08 Btu/hr respectively. The calculated second law efficiency was35.8%.
Vineet et al, 2001	A steady state, multicomponent distillation model particularly suited for fractionation of crude oil has been developed based on equilibrium stage relations.	An improved scheme of numbering the equilibrium stages when side strippers are present, was found to be advantageous with respect to computation time.

CHAPTER THREE

3.0 PROCESS SELECTION AND PROCESS DESIGN

3.1 Process Selection

There are two types of processes for crude distillation unit as shown in Figure 3.1 and Figure 3.2:

3.1.1 Two Stage Distillation

A block diagram showing the two basic units; Atmospheric Distillation Unit and Vacuum Distillation Unit is presented in Figure 3.1. The diagram also shows the feed and products of the two units.

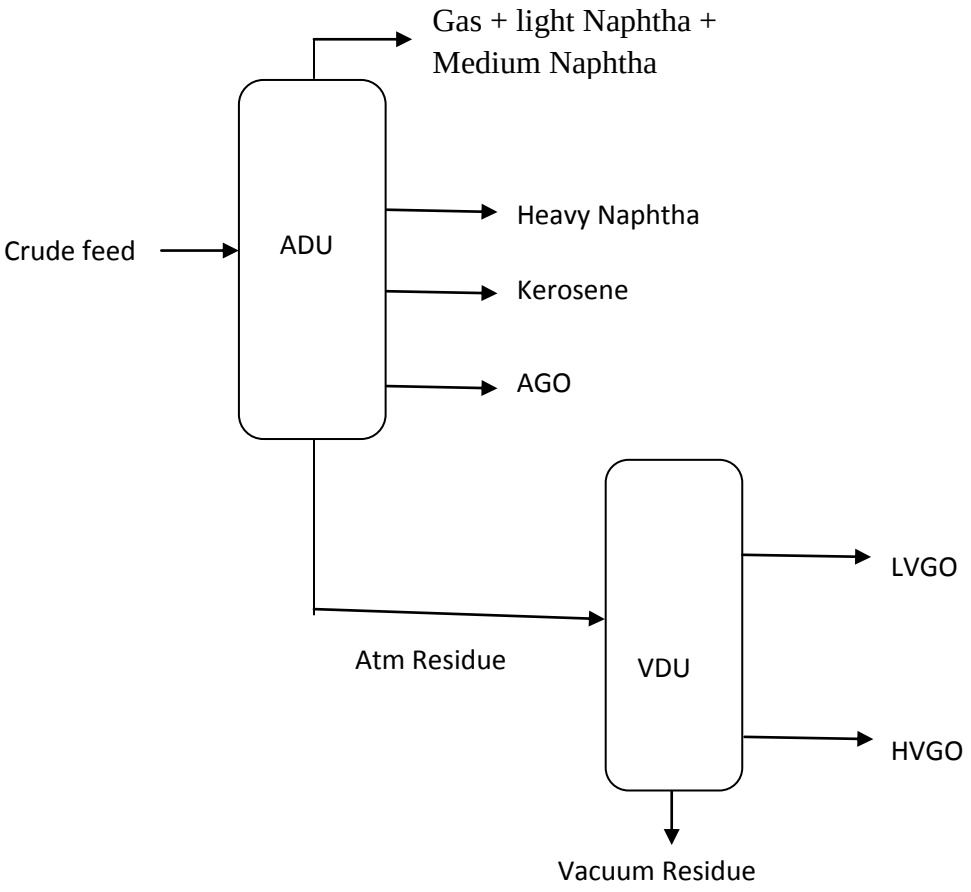


Figure 3.1: Block Diagram of Two Stage Distillation
(Aspen Plus 2004.1: Getting Started Modelling Petroleum Processes, 2005)

3.1.2 Three Stage Distillation:

It is different from the two stage distillation in that it has a pre-flash column as shown in Figure 3.2. It has advantage over the former in that more products are recovered and lesser problem of pressure build up in ADU column.

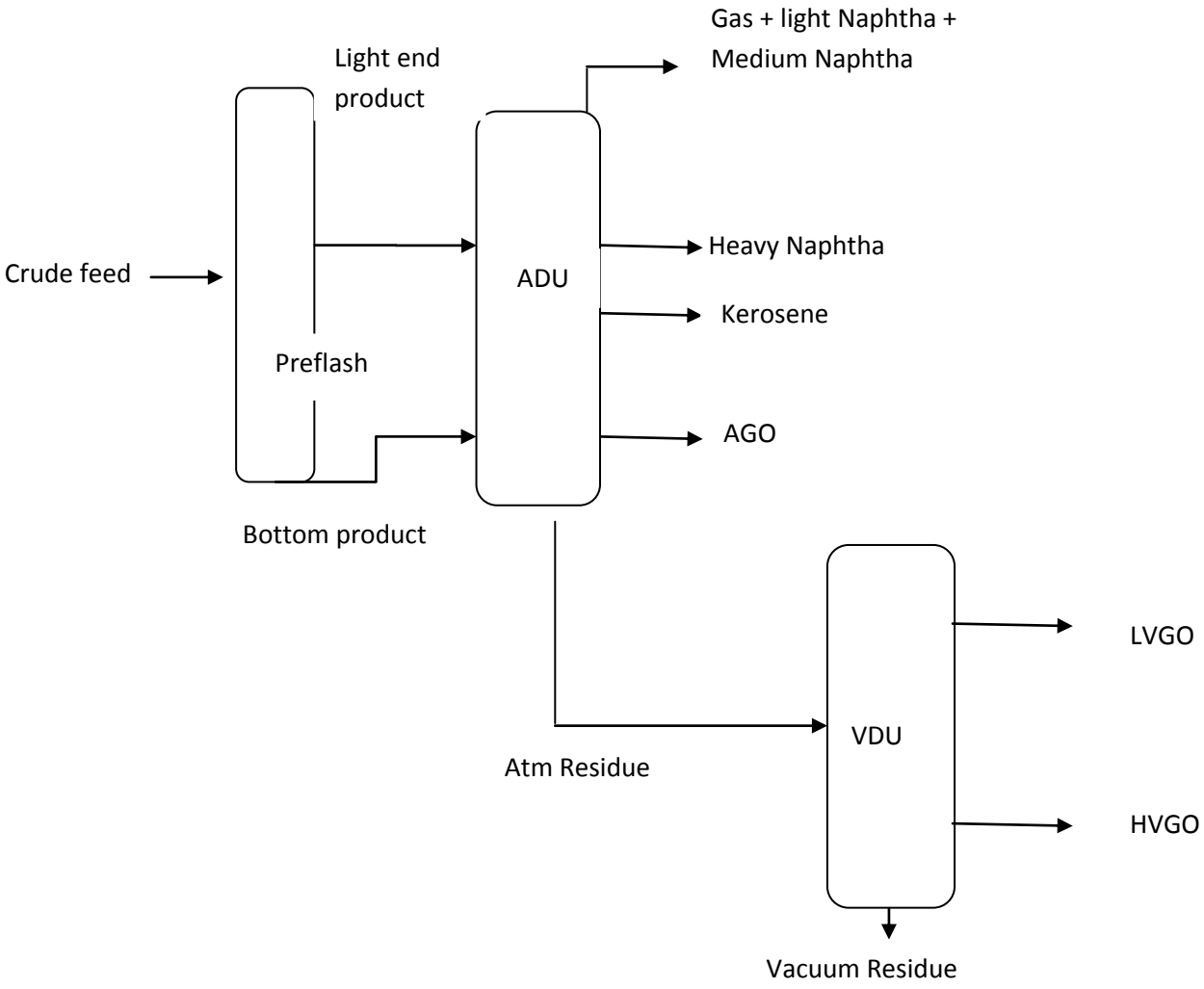


Figure 3.2: Block Diagram of Three Stage Distillation
(Aspen Plus 2004.1: Getting Started Modelling Petroleum Processes, 2005)

For this work, two stage distillation process was selected for the following reasons:

1) Lesser capital cost.
2) Lesser space requirement
3) Easier to operate and maintain.

3.2 Design Calculation for Atmospheric Distillation Column

Calculation method used all through in design of Atmospheric Distillation Column is adopted from Uppaluri (2010).

1bbl/day of Feed Crude capacity was selected to minimize capital cost at this development stage of the technology.

Escravos Crude was selected as feed crude because of it availability at nearby Kaduna Refinery and Petrochemicals located in Kaduna.

The following calculations are carried out in order to design the Atmospheric distillation column:

a) Mass balances across the ADU as well as flash zone in section 3.2.1.
b) Determination of flash zone temperature in section 3.2.2.
c) Estimation of draw off temperatures in section 3.2.3.
d) Estimation of tower top temperature in section 3.2.4.
e) Estimation of Residue Product stream temperature in section 3.2.5
f) Estimation of side stream stripper product temperatures in section 3.2.6.
g) Overall tower energy balance and estimate condenser + Bottom Pump Around + Top Pump Around duties in section 3.2.7.
h) Estimation of condenser duties in section 3.2.8.

i) Column hydraulics (I. e. estimation of overflow from top tray, verification of fractionation criteria and estimation of flash zone liquid reflux rate) in section 3.2.9 to section 3.2.12.

j) Determination of column diameter in section 3.2.13**.**

3.2.1 Material Balance

The mass balance along with relevant properties such as API, K (characterization factor) and molecular weight of various streams is carried out using refinery property as obtained from crude assay. The mass balances are desired for the overall ADU as well as the flash zone. This is due to the fact that flash zone mass balance eventually leads to energy balance across the flash zone and determination of residue product temperature. Eventually, all other sections of the main column are solvable. Figure 3.3 shows the various sections of the ADU.

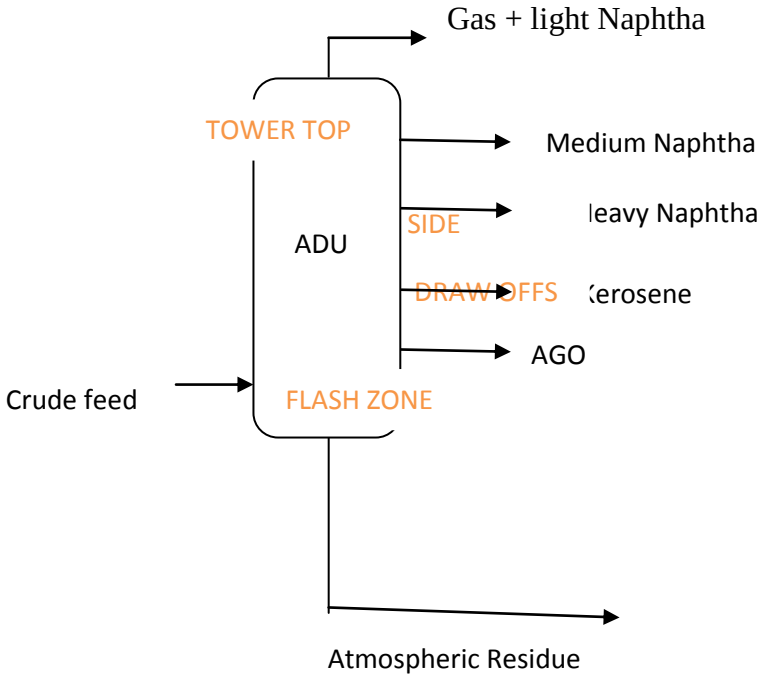


Figure 3.3: Atmospheric Distillation unit showing various sections (Uppaluri, 2010).

3.2.1.1 Mass Balance across ADU

From Escravos Crude Assay presented in Appendix A, the average properties of various products are given in Table 3.1.

Table 3.1 Average properties of products from ADU

Properties	Whole crude	Gas +light Naphtha	Medium Naphtha	Heavy Naphtha	Kerosene	LGO	Residue
Cut vol. (%)	100	5.36	15.05	11.02	13.52	20.18	34.87
S. G.	0.86	0.67	0.76	0.79	0.85	0.87	0.94
^{0}API	33.7	80	55.8	47.2	35.4	30.8	19.6
S%	0.16	0.05	0.19	0.02	0.06	0.13	0.31

For VDU, average properties of its products are presented in Table 3.2

Table 3.2 Average properties of product from VDU

Properties	Atm. Residue	LVGO	HVGO	Vacuum Residue
Cut vol. (%)	34.87	42.64	38.31	19.04
S. G.	0.94	0.91	0.94	0.99
S%	0.31	0.23	0.31	0.47

Equation of material balance

Mass flow rate, M = Volumetric flow rate(F) x 42 x 8.33 x S.G. - - 3.1

Equation for converting mass flow rate in Ib/day to Molar flow rate in mol/hr is given in Equation 3.2.

$$\text{Molar flow rate in mol/hr} = \frac{M}{(24 \times \text{Molecular weight,MW})} \quad - \quad - \quad - \quad 3.2$$

$$\text{Suphur mass flow rate, SM} = \frac{\text{Suphur content,SC} \times M}{100} \quad - \quad - \quad - \quad - \quad 3.3$$

The ADU produces five different products namely gas + Light naphtha (G+LN), Medium naphtha (MN), Heavy naphtha (HN), Kerosene, Light gas oil (LGO) and Atmospheric residue (AR).

The steady volumetric balance for the ADU is defined as

$$F_{Crude} = F_{G+LN} + F_{MN} + F_{HN} + F_K + F_{LGO} + F_{AR} \quad - \quad - \quad - \quad - \quad 3.4$$

Where, F refers to the volumetric flow rates of various streams (crude, G+LN, MN, HN, K, LGO and AR)

The mass balance for ADU is defined as

$$M_{Crude} = M_{G+LN} + M_{MN} + M_{HN} + M_K + M_{LGO} + M_{AR} \quad - \quad - \quad - \quad 3.5$$

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

Where,

F = Volumetric Flow Rate (bbl/day)

M = mass flow rate (Ibs/day)

SC = Sulfur Content (%)

SM = Sulfur Mass Flow Rate (Ibs/day)

S.G. = Specific Gravity

G + LN = Gas + Light Naphtha

MN = Medium Naphtha

HN = Heavy Naphtha

K = Kerosene

LGO = Light Gas oil

AR = Atmospheric Residue

42, 8.33, 24 and 100 are conversion constants.

Product temperature cuts are presented below,

Gas + light Naphtha (below 175^0F) (below 79^0C)

Medium Naphtha (175 – 300^0F) (79 - 149^0C)

Heavy Naphtha (300 – 400^0F) (149 - 204^0C)

Kerosene (400 – 500^0F) (204 - 260^0C)

Atm. Gas Oil (500 – 650^0F) (260 - 343^0C)

The Equations 3.1 – 3.7 were used to obtain values in mass balance for ADU.

Taking Medium Naphtha,

From Crude Assay, Medium Naphtha is 15.05% by volume of crude.

Therefore, volume flow of Medium Naphtha (bbl/day) $=\frac{\%yield \times total\ capacity}{100} = \frac{15.05 \times 1}{100}$

= 0.1505 bbl/day

Mass flow rateMedium Naphtha(Ibs/day) = 0.1505 x 42 x 8.33 x 0.76

= 40.02 Ibs/day

Sulfur flow Medium Naphtha (Ibs/day) = $\frac{40.02 \times 0.19}{100}$ = 0.076 Ibs/day

Molar rate of Medium Naphtha in mol/hr = $\frac{40.02}{24 \times 112}$ = 0.015 mol/hr

Calculation of flow carried out for Medium Naphtha above was carried on all other steams. Results of the calculations for ADU stream balances are presented in Tables 3.3.

Table 3.3: The mass balances across the ADU

Stream	Vol %	Vol. flow rate		S.G	Mass flow rate		Sulfur content (wt%)	Sulfur flow rate		MW	Molar rate Mol/hr
		bbl/day	Lit/hr		Ib/day	kg/hr		Ib/day	kg/hr		
Gas +light Naph.	5.36	0.0536	0.3551	0.67	12.56	0.2374	0.05	0.0063	1.19E-4	101	0.0052
Medium Naph.	15.05	0.1505	0.9971	0.76	40.02	0.7564	0.19	0.076	1.44E-3	112	0.015
Heavy Naph.	11.02	0.1102	0.7301	0.79	30.46	0.5757	0.02	0.0061	1.15E-4	143	0.0089
Kero-sene	13.52	0.1352	0.8957	0.85	40.21	0.7600	0.06	0.024	4.54E-4	172	0.0097
AGO	20.18	0.2018	1.3369	0.87	61.42	1.1608	0.13	0.080	1.51E-3	218	0.0117
Atm Resi-due	34.87	0.3487	2.3101	0.94	114.68	2.1674	0.31	0.356	6.73E-3	375	0.0127
Whole crude	100	1	6.6250	0.86	300.88	5.6865	0.16	0.481	9.09E-3	209	0.060

Similar mass balance carried out on VDU has result presented in Table 3.4.

Table 3.4: The mass balances across the VDU

		Vol. flow rate		S.G.	Mass flow rate		Sulfur content (wt%)	Sulfur flow rate	
Stream	Vol%	bbl/day	Lit/hr		Ibs/day	kg/hr		Ibs/day	kg/hr
LVGO	42.64	0.1487	0.985	0.91	47.34	0.895	0.23	0.109	0.0021
HVGO	38.31	0.1336	0.885	0.94	47.68	0.901	0.31	0.148	0.028
Vacuum Residue	19.04	0.0664	0.440	0.99	23.00	0.435	0.47	0.108	0.020
Atm Residue	100	0.3487	2.310	0.94	114.68	2.167	0.31	0.356	0.067

3.2.1.2 Mass Balance across Flash Zone

The crude stream fed to the flash zone in a partially vaporized state (3-5% of overflashing is usually done). The purpose of the overflashing is to enable the arrangement of the internal reflux for the control of the product quality.

%Overflash is taken as 3%

The unflashed component is the Atmospheric residue which 34.87% by volume.

So flashed volume is 100 – 34.87 = 65.13%

This means that Total vapour % = flashed volume + overflash = 65.13 + 3 = 68.13%

TBP(True Boiling Point) range on the crude = 650 – 660^{0}F is obtained from mid boiling temperature plot against cumulative volume % based on data from crude assay in Table A.2.

Overflash has property range between HGO and Atm residue

So, Molecular Weight(MW) of Overflash = $\frac{MW\ HGO + MW\ Atm\ residue}{2}$ - 3.8

$= \frac{218+375}{2}$

$= 296.5$

S. G. of Overflash = $\frac{SG\ HGO + SG\ Atm\ residue}{2}$ - - - - 3.9

$= \frac{0.87+0.94}{2}$

$= 0.905.$

Since temperature range of Products is from IBP(55^0F) to 650^0F (obtained from Crude Assay in Table A.2, property of products is an average of properties of light Naphtha, Medium Naphtha, Heavy Naphtha, Kerosene and AGO.

Molecular Weight (MW) of Products = $\frac{101+112+143+172+218}{5}$

$= 149.2$

S. G. of Products = $\frac{0.67+0.76+0.79+0.85+0.87}{5}$

$= 0.798$

Properties of Total Vapour and Residue are gotten by extrapolating plots (Figure 3.4 and 3.5) of properties vs temperature using mid boiling temperature and properties of

products obtained from Crude Assay in Table A.4.

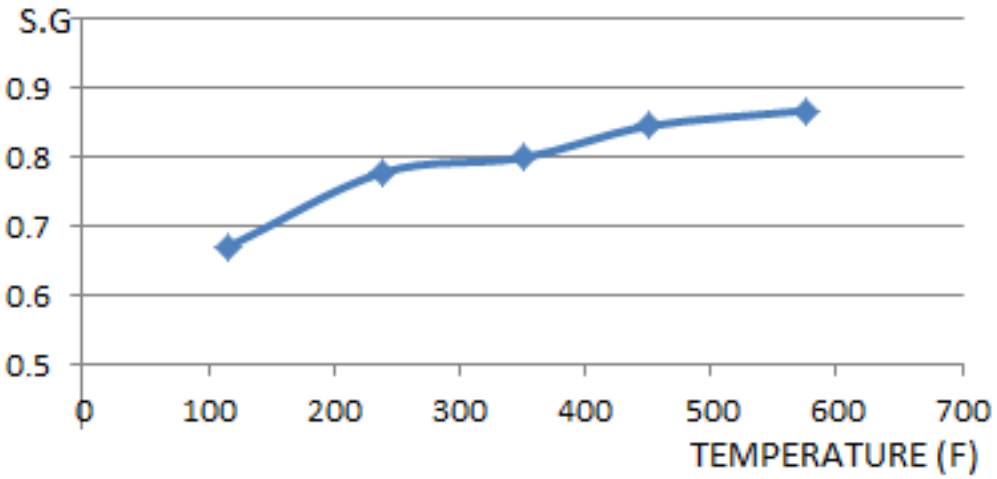


Figure 3.4: Plot of Mid BoilingTemperature(^{0}F) versus Specific Gravity(S.G)

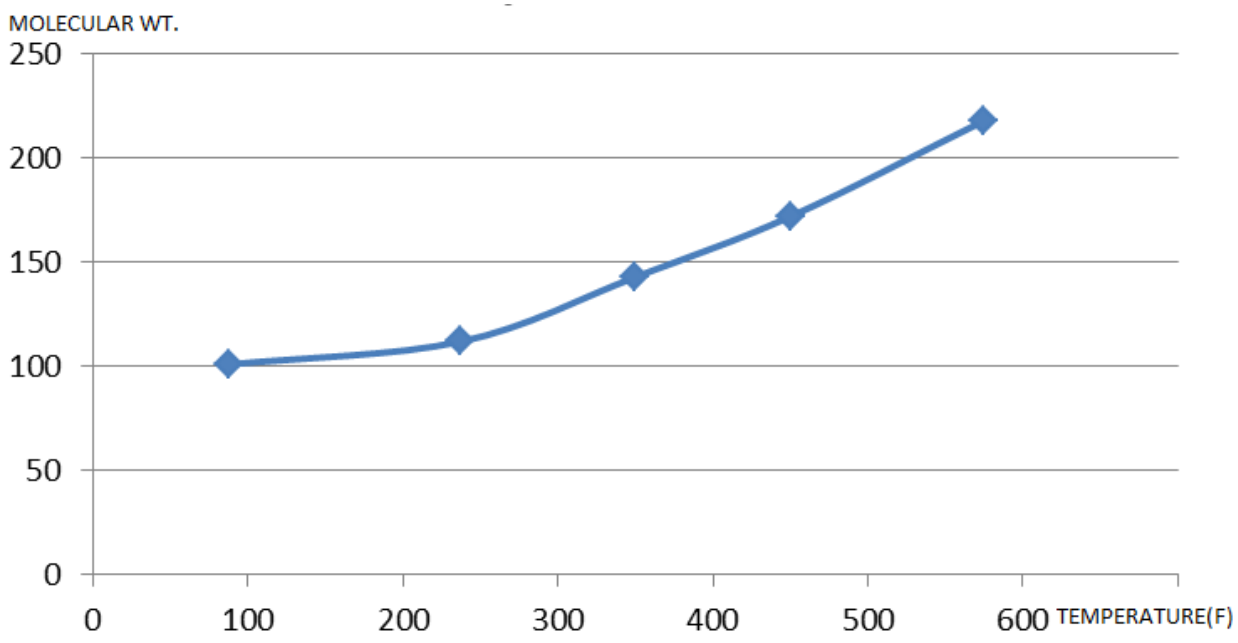


Figure 3.5: Plot of Mid Boiling Temperature vs Molecular weight

In practice boiling temperature of Residue is in surplus of 10^0F of Atmospheric Residue temperature and boiling temperature of Total vapour is in surplus of 10^0F of Products,

by extrapolating properties from Figure 3.4 and 3.5. properties of Residue was gotten by extrapolating properties of Atmospheric Residue(650^0F) by 10^0F and that of Total vapour was gotten by extrapolating properties of Products (IBP-650^0F) by 10^0F.

Residue(+10^0F) S.G = 0.95 and Molecular Weight = 378

Total Vapour S. G = 0.798 and Molecular Weight = 151.2

With the properties haven being determined, for Total Vapour

Volume flow of Total Vapour in bbl/day = $\frac{1 \times 68.13}{100}$ = 0.6813 bbl/day

By Equation 3.1

Mass flow of Total Vapour in Ib/day = 0.6813 x 42 x 8.33 x 0.798 = 190.21 Ib/day

By Equation 3.2

Molar flow of Total Vapour in mol/hr = $\frac{190.21}{24 \times 151.2}$ = 0.05 mol/hr

Similar calculation of flow for Total Vapour was carried out for other streams and results presented in Table 3.5.

Table 3.5: Mass balances across the flash zone

Stream	Range		Vol %	Volumetric flow rate		S.G.	Mass flow rate		Mol Wt	Molar rate Mol/hr
	(^{0}F)	(^{0}C)		Bbl/day	liter/hr		Ib/day	Kg/hr		
Crude	-		100	1	6.625		300.88	5.687	209	0.06
Flash Zone Products										
Overflash	650-660	343-349	3	0.03	0.199	0.905	9.50	0.180	296.5	0.0013
Products	IBP-650	IBP-343	65.13	0.6513	4.315	0.788	179.56	3.394	149.2	0.05
Total vapour	IBP-660	IBP-349	68.13	0.6813	4.514	0.798	190.21	3.595	151.2	0.05
Residue	660+	349+	31.87	0.3187	2.111	0.94	104.81	1.981	378	0.01
Total	-		100	1	6.625	0.86	300.88	5.687	209	0.06

3.2.2 Estimation of Flash Zone Temperature

The flash zone temperature is estimated using the Equilibrium Flash Vaporization (EFV) curve data of the crude for assumed overflash conditions and partial pressure of the hydrocarbons. The partial pressure concept is extremely important in mass and energy balances carried out in various sections of the ADU as steam enthalpy is a function of the partial pressure of steam that exists in the chosen zone of calculation.

EFV curve data of crude oil at 1 atm presented in Table 3.7 was obtained by Equations 3.11 -3.16 and Maxwell's correlations in Table C.14.

Slope of TBP = $S_{TBP} = \frac{T_{TBP,70} - T_{TBP.10}}{60}$ - - - - 3.11

Where $T_{TBP,70}$ = 800 and $T_{TBP,10}$ = 217 from TBP data in Table A.5. Note: $T_{TBP,10}$means True Boiling Point at Volume fraction of 10%.

Slope of TBP = $\frac{800-217}{60}$ = 9.72

Equation of DRL(Distillation Reference line)

$Y - y_{(at\ x\ =10)} = S_{TBP}\ (X - X_{(at\ x\ =10)})$ - - - - - - 3.12

$Y - 217 = 9.72(X - 10)$

$Y = 9.72X + 119.8$ - - - - - - - 3.13

Equation 3.13 is used to obtain T_{DRL} as Y for vol% 0, 10, 30, 50, 70, 90 and 100 by substituting the corresponding vol% as X.

For Vol% = 0

$T_{DRL} = 9.72(0) + 119.8$

$= 120^0F$

For Vol% = 10

$T_{DRL} = 9.72(10) + 119.8$

$= 217^0F$

T_{DRL} for other Vol% were calculated similarly and presented in Table 3.6

The slope of the FRL, S_{FRL} is found from the Maxwell's correlation (Table C.14) as $S_{FRL} = 6.531$(at $S_{TBP} = 9.72$).

From second Maxwell's correlation (Table C.14), the difference between the DRL and at 50% volume distilled was obtained as,

$\Delta t_{50(DRL - FRL)} = 40^0F$ - - - - - - - 3.14

$T_{DRL,50}= 606^0F$

Then,

$T_{FRL,50}= 606 - 40 = 566^0F$

Equation of FRL (Flash Reference line)

$Y - y_{(at\ x\ =50)} = S_{FRL}\ (X - X_{(at\ x\ =50)})$

$Y - 566 = 6.531(X - 50)$

$Y = 6.531X + 239.45$ - - - - - - - 3.16

Equation 3.16 is used to obtain T_{FRL} as Y for vol% 0, 10, 30, 50, 70, 90 and 100 by substituting the corresponding vol% as X.

For Vol% = 0

$T_{FRL} = 6.531(0) + 239.45$

$= 239^0F$

For Vol% = 10

$T_{FRL} = 6.531(10) + 239.45$

$= 305^0F$

T_{FRL} for other Vol% were calculated similarly and presented in Table 3.6

Table 3.6: Table for Crude TBP, DRL and FRL Temperature Data

Vol %	T_{TBP} (^0F)	T_{TBP} (^0C)	T_{DRL} (^0F)	T_{DRL} (^0C)	T_{FRL} (^0F)	T_{FRL} (^0C)
0	-30	-34	120	49	239	115
10	217	103	217	103	305	152
30	414	212	411	211	435	224
50	593	312	606	319	566	297
70	800	427	800	427	697	369
90	1056	569	995	535	827	442
100	1357	736	1092	589	893	478

From Maxwell's third correlation for the EFV in Table C.14, the EFV data is obtain as shown in Table 3.7.

$\Delta t_{EFV-FRL} / \Delta t_{TBP-DRL}$ - - - - - - - 3.17

Where

$\Delta t_{EFV-FRL} = T_{EFV} - T_{FRL}$ - - - - - - - 3.18

Also, $\Delta t_{TBP-DRL} = T_{TBP} - T_{DRL}$ - - - - - - 3.19

Calculation of T_{EFV}:

At Vol% = 30

$T_{TBP} = 414^0F$

$T_{DRL} = 411^0F$

$T_{FRL} = 435^0F$

$\Delta t_{TBP - DRL} = 414 - 411 = 3^0F$

$\Delta t_{EFV - FRL} / \Delta t_{TBP - DRL} = 0.34$

$(T_{EFV} - 435)/3 = 0.34$

$T_{EFV} = 437^0F$

T_{EFV} for other Vol% was done similarly and result presented in Table 3.7

Table 3.7: The EFV data for crude

Vol %	T_{EFV} (^0F)	T_{EFV} (^0C)
0	179	82
10	305	152
30	437	225
50	562	294
70	697	369
90	847	453
100	980	527

Calculation of Flash zone temperature

From literature, density of residue is taken as 1.2 Ibs/gal, Flash zone pressure = 40psia and Partial pressure of hydrocarbon=17.93psia.

From mass balance of flash zone, Residue = 105.92 Ibs/day.

Density (Ibs/gal) $= \frac{mass\ flow\ rate}{volumetric\ flow\ rate}$. - - - - - 3.20

Using Equation 3.20, converting residue flow in Ib/day to gal/hr

Residue flow rate (gal/hr) = 0.557gal/hr

Taking Steam flow rate (Ibs/hr) = 0.67 Ibs/hr

Expressing steam flow rate in Ibmol/hr

Steam flow rate (Ibmol/hr) = 0.037

From EFV data on Table 3.7

EFV 50% of the crude = 562^0F

The existing EFV data point is taken at 50% and 14.7psia. Adjusting for new partial pressure of 17.93 psia. Since EFV 50% for this pressure is on existing graph.

Slope of existing graph = 9.72 = $\frac{X - 562}{17.93 - 14.7}$

X = 593

Therefore, EFV 50% at 17.93 psia = 593^0F

Slope of the FRL = 6.531 (as previously calculated)

EFV at 17.93psia is defined using the line Y = 593 + 6.531(X – 50) where X refers to the cumulative volume%. For a value of 68.13% of X (which is the overflash value), Y = 711.4^0F which is the flash zone temperature.

3.2.3 Estimation of Draw Off Temperatures

Packie's correlation in Table 3.8 is a correlation between side stream draw off temperatures and amount of light ends stripped. These correlations are used along with relevant assumptions to estimate the draw off temperatures.

Table 3.8: Packie's correlation data to estimate the draw off temperature (Uppaluri, 2010).

0% point of cut on atmospheric TBP curve of product ^{0}F	Theoretical side draw temperature – Actual side draw temperature
197.75	9.29
301.27	24.43
300.12	24.43
500.32	66.37
599.31	95.31
686.79	122.62
743.21	142.95
798.48	162.21
851.45	183.06
897.52	201.23

Draw offs are Atm. Gas Oil, kerosene and Heavy Naphtha draw off.

Uppaluri 2010 suggested the following ratios on mass ratio basis

Steam to Atm. Gas Oil ratio = 0.5

Steam to kerosene ratio = 0.65

Steam to Heavy Naphtha ratio = 0.65

Take the following rule of thumb

Molar ratio of overflow liquid to Atm. Gas Oil product = 1.2 to 2.9

Molar ratio of overflow liquid to kerosene product = 0.9 to 1.0

Molar ratio of overflow liquid to Heavy Naphtha product = 0.9 to 1.0

Expression for calculating partial pressure is given as:

$$\text{Partial pressure} = \frac{(\text{Product vapour moles} + \text{Moles overflow}) \times \text{tray pressure}}{(\text{product vapour moles} + \text{moles steam})} \quad - \quad 3.21$$

For all three draw-off stream trays, the tray pressure is evaluated first.

The design method require product draw off trays to be taken. This parameter can be altered again and again until product requirement at the end of design calculation is met. Let us take the following draw off trays:

Atm. Gas Oil draw off tray = 5

Kerosene draw off tray = 10

Heavy Naphtha draw off tray = 16

Flash zone pressure = 40 psia

Atm. Gas Oil draw off tray is located 3 trays above the flash zone.

Average pressure drop per tray = 0.64 psia per tray

Atm. Gas Oil draw off tray pressure = 40 – 3 x 0.64 = 38.08 psia

Similarly, kerosene draw off tray pressure = 40 – 8 x 0.64 = 36.8 psia

Heavy Naphtha draw off tray pressure = 40 – 14 x 0.64 = 34.88 psia

Next, we evaluate the steam requirements in the side stream strippers.

Residue zone fresh steam flow rate (from flash zone calculations) = 0.037 Ibmol/hr

AGO zone fresh steam flow rate = 0.35 x 0.5/18 = 0.0097 Ibmol/hr

Kerosene zone fresh steam flow rate =0.24 x 0.65/18 =0.0087 Ibmol/hr

Heavy Naphtha zone fresh steam flow rate =0.19 x 0.65/18 =0.0069 Ibmol/hr

From Mass Balance Table summarized previously,

Heavy Naphtha vapour flow rate = 0.0089 Ibmol/hr

Kerosene vapour flow rate = 0.0097 Ibmol/hr

AGO vapour flow rate = 0.0117 Ibmol/hr

Hydrocarbon vapour flow rate = 0.0052 + 0.015 + 0.0089 + 0.0097 + 0.0117 = 0.0505 Ibmol/hr

i) AGO Overflow Draw off tray temperature calculation

Moles overflow = 1.5 x 0.0117 = 0.0176 Ibmol/hr

Steam flow rate = 0.037 Ibmol/hr (Only that steam that is reaching the AGO draw off tray is the steam that enters at the bottom of the main column).

Therefore, Partial Pressure of Hydrocarbons at AGO draw off tray

$$= \frac{(0.0505 + 0.0176) \times 38.1}{(0.0505 + 0.0176 + 0.037)}$$

= 24.67 psia.

Using same method used in obtaining crude EFV data as presented in Table 3.6 and 3.7, EFV data for AGO is obtained and shown in Table 3.9:

Calculation of EFV data for AGO

TBP Data was obtained from Table A.3.

$S_{TBP} = \frac{T_{TBP,70} - T_{TBP.10}}{60} = \frac{602-513}{60}$

= 1.49

Equation of DRL (Distillation Reference line)

$Y - y_{(at\ x\ =10)} = S_{TBP}\ (X - X_{(at\ x\ =10)})$

$Y = 1.49X + 498$ - - - - - - - 3.22

Where Y represent T_{DRL} and X represent Vol.%, by substituting Vol. % 0, 10, 30, 50, 70, 90, 100 in equation 3.22 DRL data was gotten and presented in Table 3.9.

S_{TBP} = 1.49 is compared with data on Maxwell first correlation on Table C.14 as shown.

$\frac{S_{FRL}}{1.49} = \frac{6.45}{9.6}$

$S_{FRL} = 1$

From second Maxwell's correlation (Table C.14), the difference between the DRL and at 50% volume distilled was obtained as,

$\Delta t_{50(DRL - FRL)} = 0^0F$

$T_{DRL,50} = 572^0F$

$T_{FRL,50} = 572 - 0$

$= 572^0F$

Equation of FRL (Flash Reference line)

$Y - y_{(at\ x\ =50)} = S_{FRL}\ (X - X_{(at\ x\ =50)})$

$Y = X + 522$ - - - - - - - 3.23

Equation 3.23 is used to obtain T_{FRL} as Y for vol% 0, 10, 30, 50, 70, 90 and 100 by substituting the corresponding vol% as X and result presented in Table 3.9.

Using Equations 3.17 – 3.19 and Maxwell's third correlation for the EFV in table C.14, the EFV data was obtained and presented in Table 3.9.

Table 3.9: EFV data for AGO

Vol %	T_{TBP} (^{0}F)	T_{TBP} (^{0}C)	T_{DRL} (^{0}F)	T_{DRL} (^{0}C)	T_{FRL} (^{0}F)	T_{FRL} (^{0}C)	T_{EFV} (^{0}F)	T_{EFV} (^{0}C)
0	500	260	498	259	522	272	521	272
10	513	267	513	267	532	278	532	278
30	538	281	542	283	552	289	551	288
50	568	298	572	300	572	300	571	299
70	602	317	602	317	592	311	592	311
90	634	334	631	333	612	322	611	322
100	650	343	646	341	622	328	621	327

From the EFV data presented in Table 3.9, the obtained IBP (initial boiling point which is at Vol% =0) for EFV curve is 521^0F.

Theoretical AGO draw off temperature is gotten by converting partial pressure of AGO to atm and then reading temperature corresponding to it and IBP from EFV data on the vertical axis of Table C.13, Theoretical AGO draw off temperature at partial pressure of 24.67psia = 1.69 atm is given as 536^0F.

From Packie's correlation (Table 3.8), for an x-axis data point of 521^0F, the y-axis point is 72^0F.

Therefore, actual AGO draw off temperature = 536 – 72

$= 464^0$F (240^0C)

NB: Other draw off temperature calculation follow same procedure as in AGO draw off.

ii) Kerosene draw off tray temperature calculation

Moles overflow = 0.9 x 0.0097 = 0.0087 Ibmol/hr

Hydrocarbon vapour flow rate = 0.0052 + 0.015 + 0.0089 + 0.0097 = 0.0388 Ibmol/hr (all HC vapours other than the AGO).

Steam flow rate =0.037 + 0.0097 = 0.0467 Ibmol/hr (this is the steam that enters at the residue zone and also in the HGO side stipper)

Therefore, partial pressure of hydrocarbons at kerosene draw off tray

$$= \frac{(0.0388 + 0.0087)\times 36.8}{(0.0388 + 0.0087 + 0.0467)}$$

= 18.8psia.

Using same method used in obtaining crude EFV data as presented in Table 3.6 and 3.7, EFV data for kerosene is obtained and shown in Table 3.10:

Calculation of EFV data for Kerosene

TBP Data was obtained from Table A.3.

$S_{TBP} = \frac{T_{TBP,70} - T_{TBP.10}}{60} = \frac{474-411}{60}$

$= 1.05$

Equation of DRL (Distillation Reference line)

$Y - y_{(at\ x=10)} = S_{TBP}\ (X - X_{(at\ x=10)})$

$Y = 1.05X + 400.5$ - - - - - - - 3.24

Where Y represent T_{DRL} and X represent Vol.%, by substituting Vol. % 0, 10, 30, 50, 70, 90, 100 in Equation 3.24 DRL data was gotten and presented in Table 3.10.

$S_{TBP} = 1.05$ is compared with data on Maxwell first correlation on Table C.14 as shown.

$\frac{S_{FRL}}{1.05} = \frac{6.45}{9.6}$

$S_{FRL} = 0.7$

From second Maxwell's correlation (Table C.14), the difference between the DRL and at 50% volume distilled was obtained as,

$\Delta t_{50(DRL - FRL)} = 0^0F$

$T_{DRL,50} = 453^0F$

$T_{FRL,50} = 453 - 0$

$= 453^0F$

Equation of FRL (Flash Reference line)

$Y - y_{(at\ x\ =50)} = S_{FRL}\ (X - X_{(at\ x\ =50)})$

$Y = 0.7X + 419$ - - - - - - - 3.25

Equation 3.25 is used to obtain T_{FRL} as Y for vol% 0, 10, 30, 50, 70, 90 and 100 by substituting the corresponding vol% as X and result presented in Table 3.10.

Using Equations 3.17 – 3.19 and Maxwell's third correlation for the EFV in Table C.14, the EFV data was obtained and presented in Table 3.10.

Table 3.10: EFV data for kerosene

Vol %	T_{TBP} (^0F)	T_{TBP} (^0C)	T_{DRL} (^0F)	T_{DRL} (^0C)	T_{FRL} (^0F)	T_{FRL} (^0C)	T_{EFV} (^0F)	T_{EFV} (^0C)
0	400	204	401	205	419	215	419	215
10	411	211	411	211	426	219	426	219
30	433	223	432	222	439	226	439	226
50	454	234	453	234	453	234	453	234
70	474	246	474	246	467	242	467	242
90	492	256	495	257	480	249	479	248
100	500	260	506	263	487	253	485	252

From the EFV data presented in Table 3.10, the obtained IBP (initial boiling point which is at Vol% =0) for EFV curve is 419^0F.

Theoretical Kerosene draw off temperature is gotten by converting Partial Pressure of kerosene to atm and then reading temperature corresponding to it and IBP from EFV data on the vertical axis of Table C.13, Theoretical Kerosenedraw off temperature at partial pressure of 18.8psia = 1.28 atm is given as 427^0F.

From Packie's correlation (Table 3.8), for an x-axis data point of 419^0F, the y-axis point is 49^0F.

Therefore, actual Kerosene draw off temperature = 427 – 49

$= 378^0$F (192^0C)

iii) Heavy Naphtha draw off tray temperature calculation

Moles overflow = 0.9 x 0.0089 = 0.0080 Ibmol/hr

Hydrocarbon vapour flow rate = 0.0052 + 0.015 + 0.0089 = 0.0291 Ibmol/hr (all Hyrocarbon vapours other than the kerosene and AGO).

Steam flow rate =0.037 + 0.0097 + 0.0087 = 0.0554 Ibmol/hr (this is the steam that enters at the Residue zone and also in the AGO and Kerosene side stipper)

Therefore, partial pressure of hydrocarbons at Heavy Naphtha draw off tray = $\frac{(0.0291 + 0.0080)\times 34.88}{(0.0291 + 0.0080 + 0.0554)}$ = 14psia.

Using same method used in obtaining crude EFV data as presented in Table 3.8, EFV data for Heavy Naphtha is obtained and shown in Table 3.11:

Calculation of EFV data for Heavy Naphtha

TBP Data was obtained from Table A.3.

$S_{TBP} = \frac{T_{TBP,70} - T_{TBP.10}}{60} = \frac{370-312}{60}$

$= 1$

Equation of DRL (Distillation Reference line)

$Y - y_{(at\ x\ =10)} = S_{TBP}\ (X - X_{(at\ x\ =10)})$

$Y = 0.97X + 302$ - - - - - - - 3.26

Where Y represent T_{DRL} and X represent Vol.%, by substituting Vol. % 0, 10, 30, 50, 70, 90, 100 in equation 3.26, DRL data was gotten and presented in Table 3.11.

$S_{TBP} = 1$ is compared with data on Maxwell first correlation on Table C.14 as shown.

$\frac{S_{FRL}}{1} = \frac{6.45}{9.6}$

$S_{FRL} = 0.7$

From second Maxwell's correlation (Table C.14), the difference between the DRL and at 50% volume distilled was obtained as,

$\Delta t_{50(DRL - FRL)} = 0^0F$

$T_{DRL,50} = 351^0F$

$T_{FRL,50} = 351 - 0$

$= 351^0F$

Equation of FRL (Flash Reference line)

$Y - y_{(at\ x\ =50)} = S_{FRL}\ (X - X_{(at\ x\ =50)})$

$Y = 0.7X + 319$ - - - - - - - 3.27

Equation 3.27 is used to obtain T_{FRL} as Y for vol% 0, 10, 30, 50, 70, 90 and 100 by substituting the corresponding vol% as X and result presented in Table 3.11.

Using Equations 3.17 – 3.19 and Maxwell's third correlation for the EFV in Table C.14, the EFV data was obtained and presented in Table 3.11.

Table 3.11: EFV data for Heavy Naphtha

Vol %	T_{TBP} (^{0}F)	T_{TBP} (^{0}C)	T_{DRL} (^{0}F)	T_{DRL} (^{0}C)	T_{FRL} (^{0}F)	T_{FRL} (^{0}C)	T_{EFV} (^{0}F)	T_{EFV} (^{0}C)
0	290	143	302	150	319	159	307	153
10	312	156	312	156	326	163	326	163
30	330	166	331	166	339	171	339	171
50	353	178	351	177	351	178	354	179
70	370	188	370	188	367	186	367	186
90	390	199	389	198	380	193	380	193
100	410	210	399	204	387	197	391	199

From the EFV data presented in Table 3.11, the obtained IBP (initial boiling point which is at Vol% =0) for EFV curve is 307^0F.

Theoretical Heavy naphtha draw off temperature is gotten by converting partial pressure of Heavy naphtha to atm and then reading temperature corresponding to it and IBP from EFV data on the vertical axis of Table C.13, Theoretical Heavy naphtha draw off temperature at partial pressure of 14psia = 0.95 atm is given as 317^0F.

From Packie's correlation (Table 3.8), for an x-axis data point of 307^0F, the y-axis point is 25^0F.

Therefore, Actual Heavy naphtha draw off temperature = 317 – 25

$= 292^0F\ (144^0C)$

3.2.4 Estimation of Tower Top Temperature

The following standards and assumptions were recommended by Uppalluri 2010.

Set the Reflux Drum Temperature and Pressure as 100^0F and 10 psig. Assume 5 psia pressure drop and hence, Tower Top Pressure = 15 psig or 29.7 psia

Assume external reflux as 1.5 times the total moles overhead product.

Vapour to top of tower is that of Gas + light Naphtha + Medium Naphtha

Vapour Flow rate at top of tower = Total Mole Overhead Product + External Reflux

$= (0.0052 + 0.015) + 1.5\ (0.0052 + 0.015)$

$= 0.0505$ Ibmol/hr

Calculations for Residue zone, AGO zone, Kerosene zone and Heavy naphtha zone fresh steam flow rate is presented in Section 3.2.3.

Steam flow rate in the tower top section = Summation of fresh steam flow rate in the Residue zone, AGO zone, Kerosene zone and Heavy naphtha zone

$= 0.037 + 0.0097 + 0.0087 + 0.0069$

$= 0.0623$ Ibmol/hr

Partial pressure of hydrocarbons in the Tower top section

$$= \frac{vapor\ flow\ at\ tower\ top \times tower\ top\ pressure}{(vapor\ flow + steam\ flw\ at\ tower\ top)}$$

$$= \frac{0.0505 \times 29.7}{(0.0505 + 0.0623)}$$

= 13.3 psia = 0.9048 atm

For determining the Tower top temperature, it is first assume a Tower top temperature of 250^0F. At this temperature, using Maxwell's Vapour Pressure correlation (Table C.13), the corresponding Vapour Pressure is determined. Result presented in Table 3.13.

Detail procedure for obtaining data presented in Table 3.13 is shown below:

From Table A.3, plotting TBP Temp at Start and TBP Temp at End against Yield at Start and Yield at End, Temp range with corresponding Cumulative Vol% for pseudocomponent data of crude was obtained and presented in Table 3.12.

Table 3.12: Psuedocomponent data for Crude

Temp range(^{0}F)	Temp range(^{0}C)	Cumulative vol(%)	Mid BP(^{0}F)	Mid BP (^{0}C)	Differential Volume(%)
60-90	16-32	0-0.8	75	24	0.8
90-110	32-43	0.8-2.1	100	38	1.3
110-140	43-60	2.1-3.1	125	52	1.0
140-160	60-71	3.1-4.2	150	66	1.1
150-170	66-77	4.2-5.8	160	71	1.6
170-220	77-104	5.8-8.4	195	91	2.6
220-250	104-121	8.4-13.8	235	113	5.4
250-280	121-138	13.8-18.1	265	129	4.3
280-320	138-160	18.1-22.1	300	149	4.0

Using Mid Boiling Point Temperature of product and their S. G. as given in Crude Assay, plot Temperature against S. G. as shown in Figure 3.5 was made.

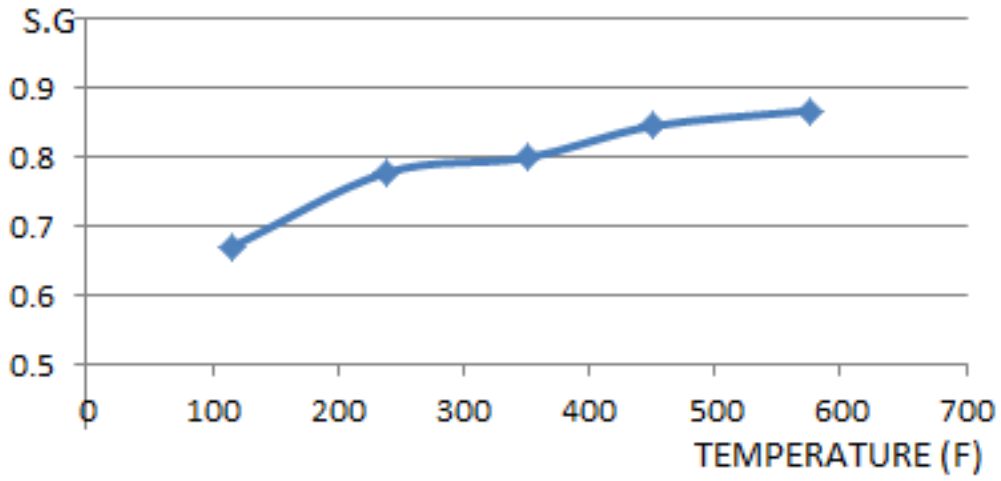


Figure 3.5: Plot of Temperature (^{0}F) against S. G.

For Mid Boiling Point 75^0F. From Figure 3.5, S. G corresponding to the Temperature is 0.639.

Weight Factor = Vol% x S.G. - - - - - - 3.28

= 3.6 x 0.639

= 2.2997

Mole fraction = $\frac{\text{Individual weight factor}}{\text{total weight factor}}$ - - - - - 3.29

= $\frac{2.2997}{73.295}$

= 0.03138

Vol% on Crude = $\frac{Differential\ volume \times 100}{Summation\ of\ differential\ volume}$

= $\frac{0.8 \times 100}{22.1}$

= 3.6%

From Maxwell Vapour pressure correlation in Table C.13.To read vapour pressure corresponding to Mid BP 75^0F, take temperature 250^0F (assumed top temperature) at vertical axis and 75^0F at horizontal axis. The vapour pressure that corresponds to these points is 12.4.

API = $(\frac{141.5}{S.G}) - 131.5$ - - - - - - - 3.30

= 90^0 API

From Table C.20, at 75^0F and 90^0API, by extrapolation Characteristic factor, K = 13.705.

$X = \frac{y}{K}$ - - - - - - - - 3.31

$= \frac{0.3138}{13.705}$

= 0.00229

Same calculation of parameters for MBP 75^0Fwas carried out for other Mid Boiling points (100, 125,….., 265, 300), to obtain Table 3.13.

Table 3.13: Vapour Pressure(Ps) data at various Crude Vol% cuts.

Mid BP, ^{0}F	Vol% on crude	S.G.	Wt Factor	Mole Fraction (y)	Ps from Maxwell	K	X=y/K
75	3.6	0.639	2.2997	0.03138	12.4	13.705	0.00229
100	5.9	0.652	3.8492	0.05252	8.56	9.461	0.00555
125	4.5	0.666	2.9966	0.04088	6	6.631	0.00616
150	5.0	0.689	3.447	0.04703	4.34	4.797	0.00980
160	7.2	0.699	5.0314	0.06865	3.8	4.200	0.0163
195	11.8	0.724	8.5408	0.1165	2.1	2.321	0.0502
235	24.4	0.747	18.2195	0.2486	1.36	1.503	0.1654
265	19.5	0.767	14.9546	0.2040	0.7	0.774	0.2636
300	18.1	0.771	13.9569	0.1904	0.52	0.575	0.3311
Total	100		73.2957	1			0.906

Taking pseudo component 9, Mid BP = 300^0F

$K_2 = Ps._{300} \times \sum X$

$K_2 = 0.52 \times 0.906$

= 0.47 atm

From Maxwell's vapor pressure table (Table C.13), component temperature 300^0F and pressure 0.47 atm corresponds to 252^0F (122^0C). This is taken as the tower temperature for future calculations.

NOTE: For preceding calculation that involves heat balance, the following important information needs to be remembered always:

a) Hydrocarbon stream's liquid or vapour enthalpy is determined from Maxwell's correlation. The Maxwell's correlations need to be interpolated for both average MEABP of the stream as well as its K value. Therefore, for any stream, it is mandatory to determine first the MEABP and K value. To determine K value, its API has to be known.

b) For all Product streams, their MEABP is assumed to be known. If not known, it is approximated to be same as its VABP.

c) For stripout streams, their MEABP is determined as an average of the MEABP values corresponding to the adjacent cuts. Same is the case for the K value as well as API.

d) For fresh steam, superheated steam data at 450 ^{o}F is taken from steam tables. For steam at any other location in the column, its saturated vapour enthalpy at the prevalent partial pressure of the steam is determined (Uppaluri, 2010).

$$\text{MEABP} = \text{VABP} = \frac{(T0 + 4\,T50 + T100)}{6} \quad - \quad - \quad - \quad - \quad 3.32$$

Where,

MEABP – Mean average boiling point

VABP – Volume average boiling point

T_0, T_{50} and T_{100} are temperatures at 0, 50 and 100% Vol cut on the TBP curve of product from the Crude Assay.

For Heavy naphtha,

$T_0 = 290^0F$

$T_{50} = 353^0F$

$T_{100} = 410^0F$

MEABP for Heavy naphtha = $\frac{(290 + 4 \times 353 + 410)}{6}$

$= 352^0F$

For Kerosene,

$T_0 = 400^0F$

$T_{50} = 454^0F$

$T_{100} = 500^0F$

MEABP for Kerosene = $\frac{(400 + 4 \times 454 + 500)}{6}$

$= 453^0F$

AGO and Residue MEABP are calculated in same manner.

MEABP for stripout from Kerosene stripper = $\frac{(\text{MEABP of Heavy naphtha} + \text{MEABP of kerosene})}{2}$ - 3.33

$= \frac{(352 + 453)}{2}$

$= 403^0F$

Other MEABP stripouts are calculated similarly.

API for stripouts were also determined by taking adjacents of product API as obtained from crude Assay.

VABP and API for streams being determined, Characterization factor, K is gotten from Table C.20 for all streams.

Mass rate of products in Ib/hr is obtained from mass balance in Table 3.1.

Haven known MEABP, K, Temperature in ^0F and phase of component (liquid or vapour), enthalpy in Btu/Ib is determined from Table C.1 – C.12 for all streams except steam stream for which its enthalpy is determined from Table C.19.

Heat (Btu/hr) = enthalpy(Btu/Ib) x mass rate (Ib/hr) - - - 3.34

For say, Residue,

Enthalpy of residue = 397.8 Btu/Ib

Mass rate of residue = 4.78 Ib/hr

Heat rate of residue = 397.8 x 4.78

= 1 901.5 Btu/hr.

3.2.5 Estimation of Residue Product Stream Temperature

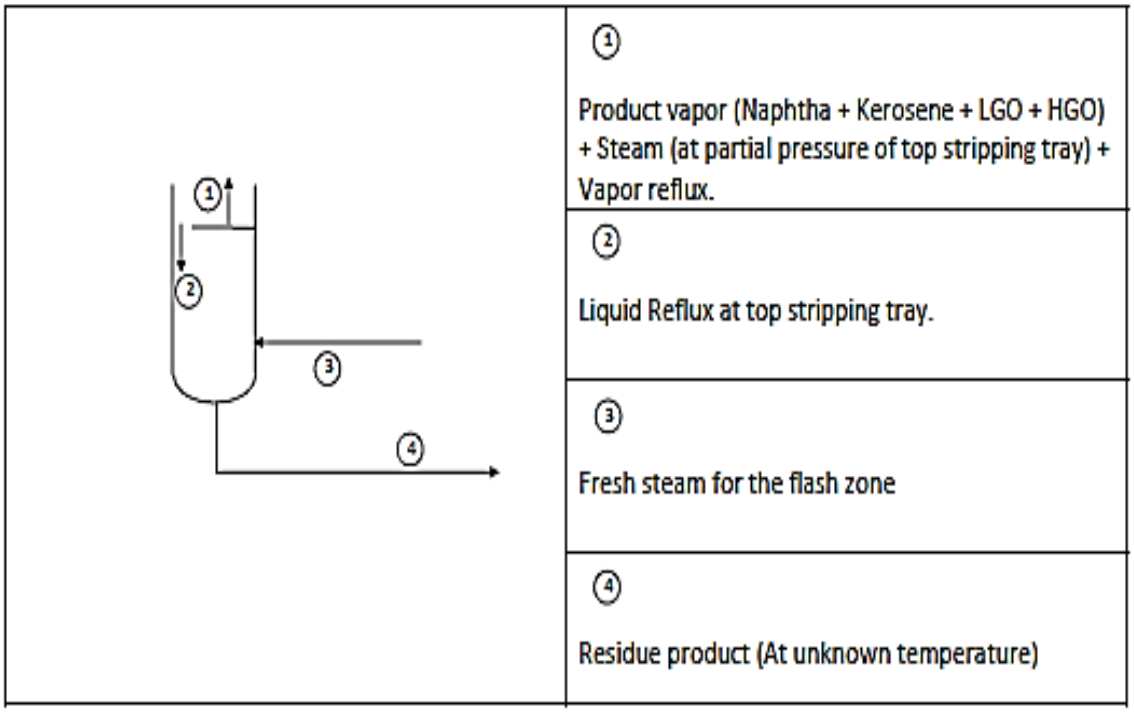


Figure 3.6: Envelope for the enthalpy balance to yield residue product temperature.

Carrying out enthalpy balance for the envelope shown in Figure 3.6, result is presented in Tables 3.14 and 3.15;

Assume for stripout + hydrocarbon vapour leaving the flash zone, their temperature is 5^0F lower than the flash zone and stripout is 0.55% of residue. (Uppaluri 2010)

The fresh steam is taken as superheated steam at 450^0F and 50 psig.

Table 3.14: Incoming streams to the Flash zone

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Residue	L	20.9	895	11.9	711.4	377.4	4.78	2.168	397.8	1901.5
Stripout	L	26.3	728	11.8	711.4	377.4	0.26	0.118	400.2	104.1
Fresh steam	V				450	232	0.67	0.304	1262.6	845.9
Total							5.71	2.590		2851.5

Table 3.15: Outing streams from the Flash zone.

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Residue	L	20.9	895	11.9	T	t	4.78	2.168	x	4.78x
Stripout	V	26.3	728	11.8	706.4	374.7	0.26	0.118	486.07	126.4
Fresh steam	V				706.4	374.7	0.67	0.304	1385	928.0
Total							5.71	2.590		1054.4 + 4.78x

Equating the incoming and outgoing heat balance total in equation 3.35,

$1054.4 + 4.78x = 2851.5$ - - - - - 3.35

$$X = \frac{(2851.5 - 1054.4)}{4.78}$$

= 376 Btu/Ib

The enthalpy of the Residue liquid product(X) = 376.0 Btu/Ib.

From Table 3.15, for liquid state, enthalpy = 367 Btu/Ib, K = 12 and MEABP = 895^0F, extrapolating from Table C.9 and C.11, corresponding temperature, T = 689.5^0F which is the Residue Product Temperature.

3.2.6 Estimation of Side Stripper Products Temperature

Consider the side stripper itself as an envelope for the heat balance. For the side stripper the incoming streams are:

a) Product draw off stream at draw off temperature that was evaluated using Packie's correlation (Table 3.8). The product draw off stream (liquid) consists of two portions namely the hydrocarbon vapor equivalent to the final product flow rate and strip out stream that enters the side stripper as liquid and leaves the side stripper as vapor (Uppaluri, 2010).

b) Fresh steam at 450^0F and 50 psig (Uppaluri, 2010).

Similarly, the outgoing streams are:

a) Side stripper product stream at unknown temperature.

b) Stripped hydrocarbon vapor whose temperature is assumed to be 5^0F lower than the draw off temperature.

c) Steam at its partial pressure and vapor temperature (Uppaluri, 2010).

In these enthalpy balances, the stripout vapor/liquid flowrate is estimated to be 5% for AGO side stripper unit and 8% for kerosene, Heavy Naphtha and Medium Naphtha side stripper units. Subsequently, API is determined to be the average of the corresponding API values of the adjacent cuts. Since in the above consolidated enthalpy balance stream only side stripper product stream enthalpy is unknown it can be determined. From the enthalpy values using Maxwell's correlation, one can determine the side

stripper product's temperature. The same procedure is applicable for Medium Naphtha, Heavy Naphtha, kerosene and AGO side strippers (Uppaluri, 2010).

The enthalpy balance table for the AGO side stripper is presented in Table 3.16 and 3.17:

Table 3.16: Incoming streams to AGO side stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Feed ex strip out	L	31.6	561	11.6	464	240	2.56	1.161	237.93	609.10
Stripout	L	33.8	508	11.6	464	240	0.13	0.059	240.58	31.28
Fresh steam	V				450	232	0.17	0.077	1262.6	214.64
Total										855.02

Table 3.17: Outgoing streamsfrom AGO side stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
AGO	L	31.6	561	11.6	T	T	2.56	1.161	x	2.56x
Stripout	V	33.8	508	11.6	459	237	0.13	0.059	347.46	45.17
Fresh steam	V(PP= 19.21psia)				459	237	0.17	0.077	1268.1	215.58
Total										260.75 + 2.56x

Solving for x by heat balance in equation 3.36,

$260.75 + 2.56x = 855.02$ - - - - - 3.36

Enthalpy of AGO product, x = 232.14 Btu/Ib

From Table 3.17, for liquid state, Enthalpy = 232.14 Btu/Ib, K = 11.6 and MEABP = 561^0F, extrapolating from Table C.8 and C.9, corresponding temperature, T = 451^0F (233^0C) which is the AGO product stream temperature.

Next, Kerosene stripper enthalpy balance is presented in Table 3.18 and 3.19.

Table 3.18: Incoming streams to Kerosene stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Feed ex strip out	L	36.0	453	11.5	378	192	1.68	0.762	193.43	324.96
Stripout	L	40.9	403	11.6	378	192	0.13	0.059	196.37	25.53
Fresh steam	V				450	232	0.16	0.073	1262.6	202.02
Total										552.51

Table 3.19: Outgoing streams from Kerosene stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Kerosene	L	36.0	453	11.5	T	T	1.68	0.762	X	1.68x
Stripout	V	40.9	403	11.6	373	189	0.13	0.059	315.20	40.98
Fresh steam	V				373	189	0.16	0.073	1226.35	196.22
Total										237.2 + 1.68x

From heat balance in Equation 3.37,

237.2 + 1.68x =552.51 - - - - - 3.37

Enthalpy of kerosene product, x = 187.7 Btu/Ib.

From Table 3.19, for liquid state, enthalpy = 187.7 Btu/Ib, K = 11.5 and MEABP = 453^{0}F, extrapolating from Table C.6 and C.8, corresponding temperature, T = 373^{0}F (189^{0}C) which is the kerosene product stream temperature.

Next, the Heavy Naphtha stripper enthalpy balance is presented in Table 3.20 and 3.21

Table 3.20: Incoming streams to Heavy Naphtha stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Feed ex strip out	L	45.7	352	11.7	292	144	1.27	0.576	148.26	188.29
stripout	L	48.15	297	11.6	292	144	0.10	0.0454	150.75	15.08
Fresh steam	V						0.12	0.0544	1262.6	151.51
Total										354.88

Table 3.21: Outgoing streams from Heavy Naphtha stripper

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Heavy Naphtha	L	45.7	352	11.7	T	t	1.27	0.576	x	1.27x
Stripout	V	48.15	297	11.6	287	142	0.10	0.0454	280.87	28.09
Fresh steam	V				287	142	0.12	0.0544	1185.41	142.25
Total										170.34 + 1.27x

From heat balance in equation 3.38,

$170.34 + 1.27x = 354.88$ - - - - - - 3.38

Enthalpy of Heavy naphtha product, x = 145.31Btu/Ib.

From Table 3.21, for liquid state, Enthalpy = 145.31 Btu/Ib, K = 11.7 and MEABP = 352^0F, extrapolating from Table C.4 and C.6, corresponding temperature, T = 286^0F (141^0C) which is the kerosene product stream temperature.

3.2.7 Total Tower Energy Balance and Total Condenser Duty Estimation

The total tower balance is carried out to consolidate the following streams:

Incoming streams:

a) Crude (Vapour + Liquid): its flow rate is determined from mass balance for both liquid and vapour streams separately.

   Properties of crude vapour and crude liquid are that of Total vapour and Residue respectively. The corresponding properties were obtained from Table 3.5 for mass rate, Section 3.2.4 and 3.2.5 for V/L, MEABP, K and Temperature. Enthalpy of crude vapour obtained by extrapolating from Table C.9 and C.11, that of crude liquid is by extrapolating from Table C.3 and C.5.

   Heat flow rate = enthalpy x mass flow rate

   Heat flow rate of crude vapour = 522.15 x 7.9254

   = 4138.25 Btu/hr

   Heat flow rate of crude liquid vapour = 180.19 x 4.3671

   = 786.91 Btu/hr

Crude Enthalpy data evaluated from first principles is presented in Table 3.22:

Table 3.22: Crude enthalpy data

				Temperature		Mass rate			Heat rate
Stream	V/L	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
Crude Vapour	V	350	11.7	711.4	377.4	7.9254	3.595	522.15	4138.25
Crude Liquid	L	895	11.9	390	198.9	4.3671	1.981	180.19	786.91
Total						12.2925	5.576		4925.16

b) Steam: Fresh steam with its total flow rate is estimated and its total enthalpy entering the CDU is estimated as shown.

Steam flow rate = 0.0623 Ibmol/hr (from Section 3.2.4)

Where 1 Ibmol = 18 Ib

Mass flow rate of steam = 0.0623 x 18

= 1.1214 Ib/hr

From Table C.19, at temperature 450^{0}F (steam inlet temperature) and 40 psia (steam inlet pressure, steam enthalpy = 1263 Btu/Ib.

Heat flow rate of steam = steam enthalpy x mass rate of steam

= 1263 x 1.1214

= 1416.33 Btu/hr

Result for steam properties is presented in Table 3.23.

Outgoing streams:

All products (Residue, AGO, Kerosene, Heavy naphtha) at their respective stream temperatures (their properties: MEABP, Temperature, Mass rate, Enthalpy and Heat rate as determined previously in Section 3.2.6 is presented in Table 3.23).

The outgoing streams for the envelope are naphtha liquid product, water liquid product and cold naphtha reflux. All these streams are assumed to be at 100^0F (Uppaluri, 2010).

The balance energy of the incoming and outgoing streams is determined as the energy removed from condenser, bottom and top pump around ($Q_C + Q_{BPA} + Q_{TPA}$) (Uppaluri, 2010).

Overall enthalpy balance is summarized in Table 3.23 from where the total condenser and pump around duties can be estimated:

Table 3.23: Overall Enthalpy balance

			Temperature		Mass rate			Heat rate
Stream	V/L	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN								
Crude					12.2925	5.576		4925.16
Steam					1.1214	0.509	1263	1416.33
Total					13.4139	6.084		6341.49
OUT								
Residue	L	11.9	689.5	365	4.78	2.168	376.0	1797.28
AGO	L	11.6	451	233	2.56	1.161	232.14	594.28
Kero	L	11.5	373	189	1.68	0.762	187.7	315.34
Heavy Naphtha	L	11.7	286	141	1.27	0.576	145.31	184.54
Light + medium Naphtha	L	12.6	100	38	2.19	0.993	51.738	113.31
Refluxes								3336.74
Total								6341.49

From Table 3.23,

Total heat flow rate IN = Total heat flow rate OUT = 6341.49 Btu/hr

Refluxes heat flow rate = Total heat flow rate OUT – summation of heat flow for Residue, AGO, Kerosene, Heavy Naphtha and Light + Medium Naphtha

= 6341.49 – (1797.28 + 594.28 + 315.54 + 184.54 + 113.31)

= 3336.74 Btu/hr

Also, $Q_C + Q_{BPA} + Q_{TPA}$ = Refluxes heat flow rate - - - 3.39

= 3336.74 Btu/hr.

3.2.8 Estimation of Condenser Duty

The condenser duty is estimated using the heat balance envelope presented in Figure 3.7.

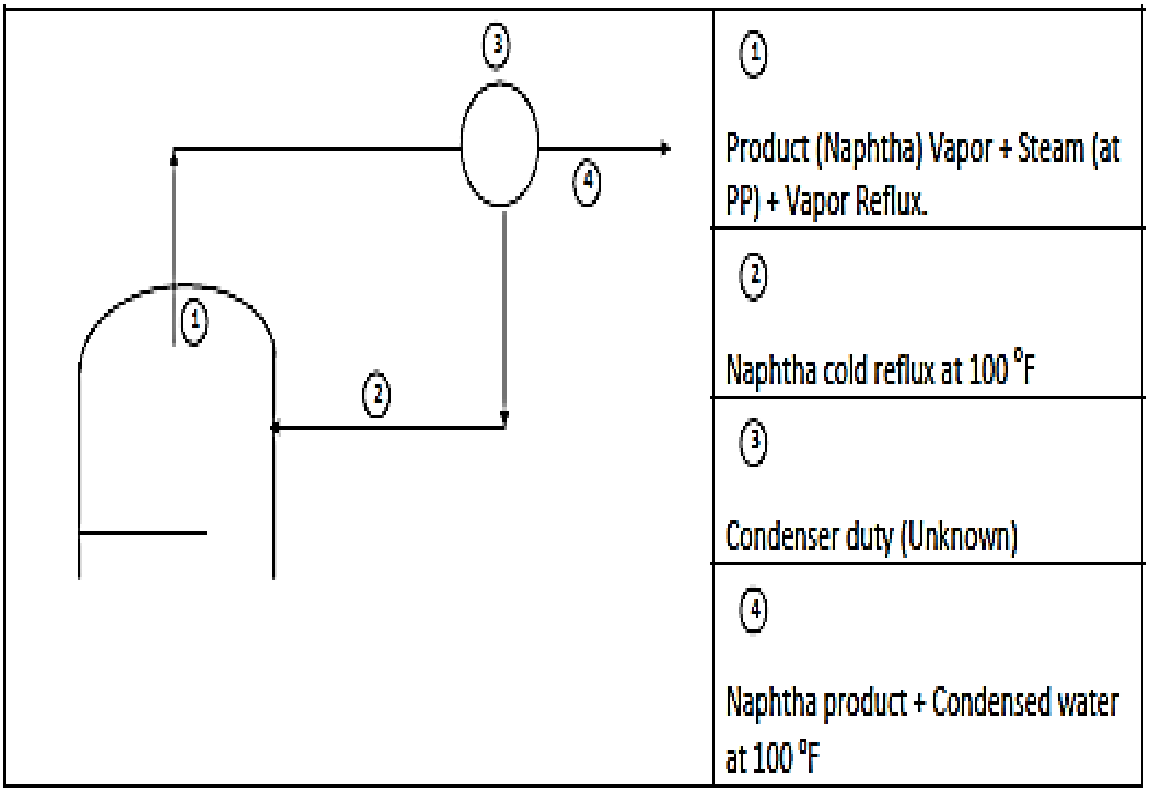


Figure 3.7: Heat balance Envelope for condenser duty estimation.

Taking 150% of Light + Medium Naphtha be refluxed (Reflux ratio of 1.5).

The enthalpy balance table for the evaluation of condenser duty is evaluated and summarized as follows in Table 3.24:

Table 3.24: Enthalpy balance table for the Evaluation of Condenser duty.

			Temperature		Mass rate			Heat rate
Stream	V/L	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN								
Light + Medium Naphtha	V	12.1	252	122	2.19	0.993	271.74	595.11
Reflux	V	12.1	252	122	3.285	1.490	271.74	892.67
Steam	V		252	122	1.1214	0.509	1151.616	1291.42
Total					5.06	2.295		2779.20
OUT								
Light + Medium Naphtha	L	12.1	100	38	2.19	0.993	51.738	113.31
Reflux	L	12.1	100	38	3.285	1.490	51.738	169.96
Water	L		100	38	1.1214	0.509	67.3	75.47
Cond. Duty								2420.46
Total					5.06	2.295		2779.20

From Table 3.24,

Total heat flow rate IN = Total heat flow rate OUT = 2779.20 Btu/hr

Condenser duty, Q_C = Total heat flow rate OUT – summation of heat flow for Water, Reflux out and Light + Medium Naphtha

= 2779.20 – (75.47 + 169.96 + 113.31)

= 2420.46 Btu/hr

Since $Q_C + Q_{TPA} + Q_{BPA}$ = 3336.74 Btu/hr - - - - 3.40

Therefore, bottom + top pump around duties ($Q_{TPA} + Q_{BPA}$) = 3336.74 – 2420.46

= 916.28 Btu/hr.

3.2.9 Estimation of Overflow from Top Tray at Reflux ratio of 4

NB: It is fair to assume that the overflow liquid is 5^0F higher than the tower top temperature and the overflow vapor is 5 – 6^0F higher than the overflow liquid. This rule of thumb will be applicable in future calculations as well that the approaching vapor and liquid are at a temperature difference of 5^0F (Uppaluri, 2010).

Let API and MEABP for Light + Medium Naphtha be average of that of Light Naphtha and Medium Naphtha while that of Overflow is average of light + Medium Naphtha and Heavy Naphtha.

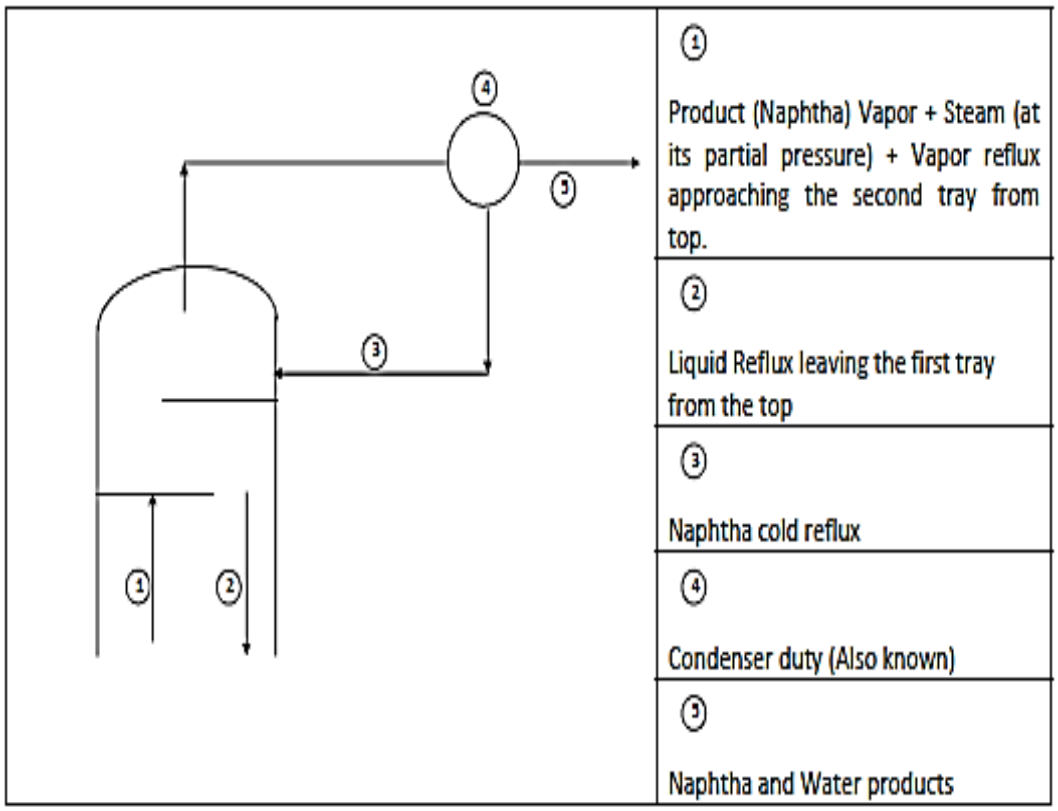


Figure 3.8: Envelope for the determination of tower top tray overflow.

The energy balance table for the envelope is evaluated and presented in Figure 3.7 is presented in Table 3.25 below:

Table 3.25: Energy balance table for Column top (Figure 3.7).

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN										
Light + Medium Naph.	V	65.4	184	12.1	262	128	2.19	0.993	277.10	606.85
O/flow	V	55.6	267	12.1	262	128	X	X	271.99	271.99x
Steam	V				262	128	1.1214	0.509	1173.51	1315.98
Total										1922.83 + 271.99x
OUT										
Light + Medium Naphtha	L	65.4	184	12.1	100	38	2.19	0.993	51.27	112.28
O/flow	L	55.6	267	12.1	257	125	X	X	140.28	140.28x
Water	L				100	38	1.1214	0.509	67.3	75.47
Cond. Duty										2420.46
Total										2608.21 + 140.28x

From the enthalpy balance,

Heat IN = Heat OUT

$1922.83 + 271.99x = 2608.46 + 140.28x$ - - - - - 3.41

Solving Equation 3.41,

$$X = \frac{(2608.46 - 1922.83)}{(271.99 - 140.28)}$$

X =5.204 Ibs/hr.

API of the overflow = 55.6 which corresponds to 6.2995 Ibs/gal.

Gallons per hour, GPH = $\frac{Ibs}{hr} \div \frac{Ibs}{gal}$ (fundamental unit expression)

Therefore, the top tray overflow in gallons per hour =$\frac{5.204}{6.2995}$ = 0.826 GPH (3.755 litre/hr). This is regarded as cold GPH (Uppaluri 2010).

Hot GPH represent the overflow volumetric flow rate in gallons per hour at its operating temperature.

3.2.10 Verification of Fractionation Criteria

Factor F is to be found as given in Equations 3.42 to 3.45:

i) LM Naphtha – Heavy Naphtha

$$F = \frac{Hot\ GPH\ leaving\ top\ tray}{LM\ Naphtha\ product\ flow\ rate\ (GPH) in\ the\ vapor} \quad - \quad - \quad - \quad - \quad 3.42$$

ii) Heavy Naphtha – Kerosene

$$F = \frac{Hot\ GPH\ leaving\ Heavy\ Naphtha\ Draw\ off\ Tray}{LM\ Naphtha{+}Heavy\ Naphtha\ product\ flow\ rate\ (GPH) in\ the\ vapor} \quad - \quad - \quad 3.43$$

iii) Kerosene – Atm Gas Oil

$$F = \frac{Hot\ GPH\ leaving\ Kerosene\ draw\ off\ tray}{LM\ Naphtha{+}Heavy\ Naphtha{+}kerosene\ product\ flow\ rate\ (GPH) in\ the\ vapor} \quad - \quad 3.44$$

iv) Atm Gas Oil – Kerosene

$$F = \frac{Hot\ GPH\ leaving\ Atm\ Gas\ Oil\ draw\ off\ tray}{LM\ Naphtha{+}Heavy\ Naphtha{+}kerosene{+}Atm\ Gas\ Oil\ product\ flow\ rate\ (GPH) in\ the\ vapor} \quad 3.45$$

Where,

Hot GPH = cold GPH $\frac{Cold\ SG}{Hot\ SG}$ - - - - - - 3.46

The number of trays in various sections is taken as follows for various combinations:

i) LM Naphtha- Heavy Naphtha: N=5(No pump arounds, therefore, all trays included in the evaluation)

ii) Heavy Naphtha-Kerosene:N=5(2 pump around trays get a credit of only 1 tray)

iii) Kerosene-Atm Gas Oil: N=4(2 pump around trays get a credit of only 1 tray)

iv) Atm Gas Oil-Residue: Criteria not required. This is due to the fact that usually ASTM gaps are defined for other combinations and the final TBP temperature of the residue is usually specified (Uppaluri, 2010).

ASTM Gaps are usually defined as per the following product specifications for the CDU

i) LM Naphtha-Heavy Naphtha: 25^0F

ii) Heavy Naphtha-Kerosene: -10^0F

iii) Kerosene-Atm Gas oil:G: -25^0F

The fractionation criteria is defined as a correlation between Reflux ratio, N, G and $\Delta T_{50\%}$ TBP between the adjacent cuts. The fractionation criteria correlation data are presented in Table C.16 and C.17 in Appendix for steam stripping conductions in the ADU (Uppaluri, 2010).

The following should be noted about reflux ratio:

a) A wrong choice of reflux ratio may provide negative flow rates. This is unacceptable.

b) Even if one chooses an appropriately high value of reflux ratio, positive reflux flow rate values may be obtained. However, fractionation criteria needs to be satisfied.

c) Very high values of reflux ratio enhances the column diameters and contributes to the cost (Uppaluri, 2010).

Verification of Fractionation criteria and determination of the optimal reflux ratio that provides all criteria (positive flow rates and ASTM gap based fractionation criteria) satisfied.

From section 3.2.9, for condenser duty = 2420.46 Btu/hr, overflow volumetric flow rate at 60^0F (cold GPH) was obtained as 0.826 GPH.

First, we apply the fractionation criteria for LM Naphtha-Heavy Naphtha combination:

From Table 3.1, Specific gravity of Light + Medium Naphtha (Cold S.G) $=\frac{0.67+0.76}{2} =$ 0.72(at 60^0F)

Specific gravity of the stream at tower top temperature (252^0F) (Hot S.G) = 0.6154 (from Table C.15)

Hot GPH =cold GPH$\frac{Cold\ SG}{Hot\ SG}$ $=\frac{0.826 \times 0.72}{0.6154}$

= 0.9645 GPH

From Table 3.3

Cold Light + Medium Naphtha product rate = 0.3551 + 0.9971

= 1.3522 litre/hr = 0.3476 GPH

Factor F = $\frac{\text{Hot GPH}}{\text{Ligth+ Medium Naphtha product rate}}$ - - - - - 3.47

= $\frac{0.9645}{0.3476}$ = 2.775

N (number of trays) = 5

Ordinate on fractionation criteria correlation = 2.775 x 5 = 13.87

From product TBP, converting TBP to ASTM. $\Delta T_{50\%}$ = 355.49 – 200.12 = 155.37^{0}F

From fractionation criteria correlation data in Table C.16, ASTM gap = 22.5^{0}F

Desired minimum ASTM gap = 25^{0}F (Uppaluri, 2010)

Since obtained ASTM gap is less than desired ASTM gap, the fractionation criteria is concluded to be unsatisfied.

Higher reflux ratios are to be selected.

Reflux ratio of 4 taken.

Calculation of condenser and pump around duty is to be recalculated with this new reflux rate taken into consideration and the fractionation criteria verified again. Higher reflux is still selected until fractionation criteria are verified. In other words, selection of reflux is an iteration that ends when fractionation criteria are met.

The enthalpy balance table for the evaluation of condenser duty is summarized in Table 3.26:

Table 3.26: Enthalpy balance for Evaluation of Condenser duty at Reflux ratio of 4.

			Temperature		Mass rate			Heat rate
Stream	V/L	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN								
Light + Medium Naphtha	V	12.1	252	122	2.19	0.993	271.74	595.11
Reflux	V	12.1	252	122	7	3.175	271.74	1902.18
Steam	V		252	122	1.1214	0.509	1151.616	1291.42
Total					5.06	2.295		3788.71
OUT								
Light + Medium Naphtha	L	12.1	100	38	2.19		51.738	113.31
Reflux	L	12.1	100	38	7	3.175	51.738	362.17
Water	L		100	38	1.1214	0.509	67.3	75.47
Cond. Duty								3237.76
Total					5.06	2.295		3788.71

From Table 3.26, condenser duty = 3237.76 Btu/hr.

Therefore, bottom + top pump around duties = 3336.74 – 3237.76 = 98.98 Btu/hr.

3.2.11 Estimation of Overflow from Top Tray at Reflux ratio of 4

NOTE: It is fair to assume that the overflow liquid is 5^0F higher than the tower top temperature and the overflow vapor is $5 – 6^0$F higher than the overflow liquid. This rule of thumb will be applicable in future calculations as well that the approaching vapor and liquid are at a temperature difference of 5^0F (Uppaluri, 2010).

Let API and MEABP for Light + Medium Naphtha be average of that of Light Naphtha and Medium Naphtha while that of O/flow is average of light + Medium Naphtha and Heavy Naphtha.

The energy balance table for the envelope presented in Figure 3.7 is presented below:

Table 3.27: Energy balance for Overflow Determination at Reflux ratio of 4.

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN										
Light + Medium Naphtha	V	65.4	184	12.1	262	128	2.19	0.993	277.10	606.85
O/flow	V	55.6	267	12.1	262	128	X	X	271.99	271.99x
Steam	V				262	128	1.1214	0.509	1173.51	1315.98
Total										1922.83 + 271.99x
OUT										
Light + Medium Naph.	L	65.4	184	12.1	100	38	2.19	0.993	51.27	112.28
O/flow	L	55.6	267	12.1	257	125	X	X	140.28	140.28x
Water	L				100	38	1.1214	0.509	67.3	75.47
Cond. Duty										3237.76
Total										2608.21 + 140.28x

From the enthalpy balance,

Heat IN = Heat OUT - - - - - - - 3.48

$1922.83 + 271.99x = 2608.21 + 140.28x$ - - - - - 3.49

we get x =11.41 Ibs/hr.

API of the overflow = 55.6 which corresponds to 6.2995 Ibs/gal.

Therefore, the top tray overflow in gallons per hour = 11.41/6.2995 = 1.81 GPH (8.23 litre/hr). This is regarded as cold GPH (Uppaluri, 2010).

Verification of Fractionation Criteria and determination of the Optimal reflux ratio that provides all criteria (positive flow rates and ASTM gap based fractionation criteria) satisfied.

From previous calculation,

Condenser duty = 3237.76 Btu/hr

Pump around duty is negligible.

First, we apply the fractionation criteria for LM Naphtha-Heavy Naphtha combination:

Also previously, we obtained the cold GPH of the liquid reflux as 1.81 GPH.

Specific gravity of the stream at tower top temperature (252^0F) = 0.6154

From Mass balance table, SG of LM Naphtha = 0.7186(at 60^0F)

Hot GPH = 1.81 x 0.7186/0.6154 = 2.115 GPH (9.61 litre/hr)

Cold LM Naphtha product GPH = 0.3476

Factor F = Hot GPH/LM Naphtha product rate

= 2.115/0.3476 = 6.085

N = 5

Ordinate on fractionation criteria correlation = 6.085 x 5 = 30.42

From product TBP, converting TBP to ASTM. We get $\Delta T_{50\%}$ = 355.49 – 200.12 = 155.37^0F

From fractionation criteria correlation data in Table C.16, ASTM gap = 38.47^0F

Desired minimum ASTM gap = 25^0F (Uppaluri, 2010)

Since obtained ASTM gap is greater than desired ASTM gap, the fractionation criteria is concluded to be satisfied.

We next carry out the energy balances for the energy balance envelope presented in Figure 3.7. Summary of the energy balance table is presented in Table 3.28 below:

Table 3.28: Energy balance for Flow analysis at Reflux ratio of 4.

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN										
Crude	V + L						12.293	5.576		4925.16
Steam	V				450	232	1.246	0.565	1262.6	1573.20
Tray 34 overflow	L	45.7	404	11.9	301	149	X	X	280.33	280.33X
Total							13.539	6.141		6498.36 + 280.33X
Out										
O/F to tray 34	V	45.7	404	11.9	306	152	X	X	283.04	283.04X
Vapour to tray 34	V	55.6	297	12.1	306	152	3.273	1.485	292.05	955.88
Steam to tray 34	V				306	152	1.246	0.565	1194.46	1488.30
Kerosene	L	36.0		11.5	373	189	1.68	0.762	187.7	315.34
Atm Gas Oil	L	31.6		11.6	451	233	2.56	1.161	232.14	594.28
Residue	L	20.9		11.9	689.5	365	4.78	2.168	376	1797.28
Total							13.539	6.141		5151.08 + 283.04X

Solving for X from Table 3.28,

$6498.36 + 280.33X = 5151.08 + 283.04X$ - - - - - 3.50

From Equation 3.50

X = 497.15 Ib/hr. which corresponds to cold GPH = 74.74 GPH (339.8 litre/hr).

Flow rate is positive, so column mass balance requirements are met.

We proceed towards applying fractionation criteria based on correlation data for side to side products

SG at 306^0F = 0.634

Hot GPH = 74.74 x 0.756/0.634 = 89.12

LM Naphtha + Heavy Naphtha product flow rate = 0.364 + 0.191 = 0.555

F = 89.12/0.555 = 160.57

N = 11

Ordinate = 160.57 x 5 = 802.87

From product TBP data, T50% TBP = 452.4 – 355.49 = 96.9^0F.

ASTM gap from Table C.17 = 315.4^0F

Since the obtained ASTM gap is higher than the desired ASTM gap, we accept this value. Since both flow rates and fractionation criteria are satisfied, Reflux ratio of 4 is satisfactory.

3.2.12 Estimation of Flash Zone Liquid Reflux Rate

The flash zone liquid reflux rate is determined by conducting the energy balance across the heat balance envelope taken from below the Atm Gas Oil draw off tray and for the flash zone (Uppaluri, 2010).

The heat balance taken for the heat balance envelope taken as Figure 3.5 is presented in Table 3.29 below:

Table 3.29: Heat balance for Estimation of Flash zone Liquid reflux rate

					Temperature		Mass rate			Heat rate
Stream	V/L	API	MEABP	K	^{0}F	^{0}C	Ib/hr	kg/hr	Btu/Ib	Btu/hr
IN										
Crude	V+L						12.29	5.575		4925.16
Steam	V				450	232	0.666	0.302	1262.6	840.89
Liquid Reflux	L	30.8		11.6	451	233	X	X	231.68	231.68X
Total										5766.05+ 231.68X
OUT										
Product vapor	V	47.2	404	12.1	456	236	7.51	3.406	254.48	1911.14
Steam in vapor	V				456	236	0.666	0.302	1265.91	843.10
Reflux vapor	V	30.8	561	11.6	456	236	X	X	348.22	348.22X
Residue	L	19.6		11.9	689.5	365	4.78	2.168		1797.28
Total										2754.24+ 348.22X

From energy balance in Table 3.29,

5766.05+ 231.68X = 2754.24+ 348.22X - - - - - 3.51

Solving Equation 3.51

x = 25.84 Ib/hr

Atm Gas Oil SG of 0.8718 which corresponds to 7.26 Ib/gal.

In gallon per hour = 25.84/7.26 = 3.56 GPH (16.18 litre/hr).

3.2.13 Estimation of Column Diameters

Let top section diameter be taken as reference diameter. Calculation of diameter is as follows and results presented in Table 3.30.

Top tray overflow = Reflux Liquid = x = 11.41 Ib/hr (from Table 3.27)

Total tower top flow (Total G) = Top tray overflow + steam flow + Light + Medium Naphtha flow = 11.41 + 1.1214 + 2.19 (from Table 3.27)

= 14.7214 Ib/hr.

Flooding value, Gf = $Kf\sqrt{(\rho V)(\rho L - \rho V)}$ - - - - - 3.51

Where,

ρV = Vapour density = 0.193156 Ib/ft^3

ρL = Liquid density = 44.5486 Ib/ft^3

Kf is a constant called Flooding factor determined from Table C.18. Taking tray spacing of 10 inches and on the sieve & valve flood line column,

Kf = 390

Gf = $390\sqrt{(0.193156)(44.5486 - 0.193156)}$

= 904.7

Maximum permissible value of Gf during operation is at 80 % flooding value.

Permissible Gf = 80% Gf = 0.8 x 904.7

= 723.76

Vapour area, A = $\frac{Total\ G}{Permissible\ Gf}$ - - - - - - 3.52

$$= \frac{14.7214}{723.76}$$

$$= 0.0203 \text{ ft}^2$$

Cfs of liquid flow = $\frac{Reflux\ liquid}{\rho L \times 3600}$ - - - - - - 3.53

$$= \frac{11.41}{44.5486 \times 3600}$$

$$= 0.000071 \text{ ft}^3/\text{s}$$

Downcomer area, Adc = $\frac{cfs}{Downcomer\ liquid\ velocity}$ - - - - 3.54

$$= \frac{0.000071}{0.02}$$

$$= 0.00356 \text{ ft}^2$$

Take an additional 20 % area as Waste area

Waste area, Aw = 20% Vapour area = 0.2 x 0.0203

$$= 0.00407 \text{ ft}^2$$

The Total area of the tray is evaluated using the expression

Total area = Vapour area + downcomer area x 2 + Waste area - - 3.55

$$= 0.0203 + 0.00356 \text{ x } 2 + 0.00407$$

$$= 0.0315 \text{ ft}^2$$

Calculated the diameter, D of the tray using the expression

D = $\sqrt{\frac{4 \times Total\ Area}{\pi}}$ - - - - - - - - 3.56

$$= \sqrt{\frac{4 \times 0.315}{\pi}}$$

= 0.2 ft

Table 3.30: Tower top sectionDiameter determination table

Reflux Liquid=	11.41 Ib/hr	5.175 kg/hr
Total G=	14.7214 Ib/hr	6.678 kg/hr
ℓV=	0.193156 Ib/ft^3	3.096 kg/m^3
SG of Liquid=	5.9557 Ib/gal	0.594 kg/litre
ℓL=	44.5486 Ib/ft^3	714.037 kg/m^3
Gf=	904.7	
Plate Area, A=	0.0203 ft^2	0.0019 m^2
Downcomer liquid velocity=0.02 ft/s		
Cfs of liquid flow	0.000071ft^3/s	1.98E-6 m^3/s
Aw	0.00407 ft^2	0.00038 m^2
Adc	0.00356 ft^2	0.00033 m^2
2 downcomers	0.00712 ft^2	0.0066 m^2
At	0.0315 ft^2	0.00293 m^2
Column diameter, D	0.20 ft	0.061 m

3.2.14 ASPEN Plus Simulation of ADU and VDU

Procedure for Aspen Plus Simulation involve drawing flowsheet for process, entering input parameters for the units, running simulation and comparing output of simulation with desired output. The input parameters are re- specified where simulation output does not met desired output. Input parameters that give desired simulation output are accepted for process design and operation.

For the Distillation Units (ADU and VDU):

Input parameters are Crude Assay, crude input conditions (temperature, flow rate and pressure), steam input conditions (temperature, flow rate and pressure), column specifications, stripper specifications and heat exchanger specifications.

These input parameters are specified based on literatures consulted and results gotten from manual calculations earlier done in Section 3.2.1 to 3.2.14.

Output parameters are product yield and conditions (temperature, flow rate and pressure) which are results of simulation run. The product yield and conditions from simulation is compared with theoretical product yield as shown in Crude Assay (Table A.1 to A.4). if the variation is unacceptable, column, stripper and heat exchanger specifications are adjusted until an acceptable variance is gotten.

The input specifications that give a narrow and acceptable variance of simulation output form crude assay output is accepted as design and operating parameters for the ADU and VDU.

Figure E.1 – E.42 present Aspen Plus Simulation User Interfaces showing input and output parameters for ADU.

Figure F.1 – F.27 present Aspen Plus Simulation User Interfaces showing input and output parameters for VDU.

3.3 Summary of Process Design Results

3.3.1 Summary of Design Result (Manual Calculation) for ADU

Product stream Mass flow rates and Temperatures obtained from manual calculation for ADU are shown on Table 3.31.

Table 3.31:Product stream Results for ADU (Manual)

Stream	Mass flow rate		Product temperature	
	Ib/hr	kg/hr	^{0}F	^{0}C
Residue	4.78	2.17	689	365
AGO	2.55	1.16	451	233
Kerosene	1.68	0.76	373	189
Heavy Naphtha	1.27	0.58	286	141
Light + medium Naphtha	2.19	0.99	100	38

Other results obtained from manual calculation for ADU are as follow:

Condenser duty

Condenser duty = 3237.76 Btu/hr

Reflux ratio = 4

ADU Design Result

No of trays = 20

Tray spacing = 10 in (25.4 cm)

Column diameter = 0.061 m

Column height = 20 x 10 = 200 in (5.08 m)

Pressure at top of tower = 28.5 psia

Flash zone temperature = 711^0F (377^0C)

Tower top temperature = 252^0F (122^0C)

Feed stage = 19

AGO drawoff stage = 16

Kerosene drawoff stage = 11

Heavy Naphtha drawoff stage = 5

3.3.2 Summary of Design Result from ASPEN Plus for ADU (Tray spacing = 10 in)

Table 3.32 shows the product stream Mass flow rate and Temperature from ASPEN Plus for ADU with Tray spacing = 10 in.

Table 3.32:Product stream Results for ADU (Aspen Plus, TS = 10in)

Stream	Mass flow rate Ib/hr	kg/hr	Product temp. ^{0}F	^{0}C
Residue	4.8	2.18	563.8	295
AGO	2.5	1.13	491.4	255
Kerosene	1.5	0.68	375.0	191
Heavy Naphtha	1.2	0.54	278.8	137
Light + medium Naphtha	2.2	1.00	127.6	53

Other results obtained from Aspen Plus Simulation for ADU with Tray spacing = 10 inches are as follow:

Condenser duty

Condenser duty = 3426.70 Btu/hr

Furnace duty = 4370.49 Btu/hr or 1101.36 kcal/hr

Furnace temperature = 609^0F

Reflux ratio = 4.18

ADU Design Result

No of trays = 20

Column diameter = 0.104 ft(0.032 m)

Side weir length = 0.075 ft (2.3 cm)

Column height = 20 x 10 = 200 in (5.08 m)

Pressure at top of tower = 15 psia

Feed stage = 18

AGO drawoff stage = 17

Kerosene drawoff stage = 13

Heavy Naphtha drawoff stage = 6

3.3.3 Summary of Design Result from ASPEN Plus for tray spacing = 6 in

Table 3.33 shows the product stream Mass flow rate and Temperature from ASPEN Plus for ADU with Tray spacing = 6 in.

Table 3.33:Product stream Results for ADU(Aspen Plus, TS = 6in)

Stream	Mass flow rate Ib/hrkg/hr		Product temp. ^{0}F	^{0}C
Residue	4.8	2.18	563.8	295
AGO	2.5	1.13	491.4	255
Kerosene	1.5	0.68	375.0	191
Heavy Naphtha	1.2	0.54	278.8	137
Light + medium Naphtha	2.2	1.00	127.6	53

Other results obtained from Aspen Plus Simulation for ADU with Tray spacing = 6 inches are as follow:

Condenser duty

Condenser duty = 3426.82 Btu/hr

Furnace duty = 4370.540 Btu/hr or 1101.37 kcal/hr

Furnace temperature = 609^{0}F (321^{0}C)

Reflux ratio = 4.19

Stripper diameter sizes obtained from Aspen Plus Simulation for Tray spacing 6 inches is shown in Table 3.34.

Table 3.34: Stripper size data

Stripper	Column Diameter, cm
1	1.48
2	1.67
3	1.78

ADU Design Result

No of trays = 20

Column diameter =0.246 ft(0.075 m)

Side weir length = 0.178 ft (0.054 m)

Column height = 20 x 6 = 120 in (3.048 m)

Pressure at top of tower = 15 psia

Feed stage = 18

AGO drawoff stage = 17

Kerosene drawoff stage = 13

Heavy Naphtha drawoff stage = 6

Hole diameter = 2 mm

Sieve hole area to active area = 0.12

Wier height = 1.5 in (3.81 cm)

Results for Aspen plus simulation for tray spacing of 6 inches was accepted because simulation product output met requirement and the optimum size obtained from the simulation, so its specification was being used for equipment design.

3.3.4 Design of Vacuum Distillation Unit (VDU)

The design was done using Aspen plus with information from crude Assay and VDU specifications obtained from literature. The design bases are as follows:

a. Feedstock = 650^0F^+ (343^0C^+) atmospheric residue from 33.7^0API Escravos crude. This corresponds to 34.87 vol. percent of the whole crude.

b. Feed rate = 0.3487 bbl/day (2.31 litre/hr) at 20.9^0API.

c. Over-flash = 0.6 vol percent of vacuum tower feed.

d. 5 vol percent of LVGO as off gas.

e. 0.4 inch weir height.

f. Steam supplied at 60 psia and 450^0F (232^0C).

Table 3.35 shows the mass balance for Vacuum Distillation Unit and Table 3.36 shows the TBP curve data for Atmospheric residue.

Table 3.35: Material Balance for VDU

Stream	Vol%	Volumetric flow bbl/day	litre/hr	S.G.	Mass flow rate Ibs/day	kg/hr	Sulfur content (wt%)	Sulfur flow Ibs/day	kg/hr
LVGO	42.64	0.1487	0.985	0.91	47.34	0.895	0.23	0.109	0.0021
HVGO	38.31	0.1336	0.885	0.94	47.68	0.901	0.31	0.148	0.0028
Vacuum Residue	19.04	0.0664	0.440	0.99	23.00	0.435	0.47	0.108	0.0020
Atm Residue	100	0.3487	2.310	0.94	114.68	2.167	0.31	0.356	0.0067

Table 3.36: TBP curve data for Atmospheric residue

Vol. %	TBP, ^{0}F	^{0}C
5	665	352
10	683	362
30	761	405
50	840	449
70	943	506
90	1161	627
95	1292	700

SUMMARY OF DESIGN RESULT (from Aspen plus) for VDU

Table 3.37 shows the product stream mass flow rate and temperature from ASPEN Plus for VDU with Tray spacing = 6 inches.

Table 3.37:Product stream Results for VDU

Stream	Mass flow rate		Product temperature	
	Ib/hr	kg/hr	^{0}F	^{0}C
Vacuum Residue	0.8	0.36	555	291
HVGO	1.7	0.77	549	287
LVGO	2.2	1.00	285	141
Off gas	0.6	0.27	142	61

Other results obtained from Aspen Plus Simulation for VDU with Tray spacing = 6 inches are as follow:

VDU Design Result

No of trays = 9

Column diameter =0.325 ft (9.9 cm)

Side weir length = 7.2 cm

Column height = 9 x 6 = 54 in (1.37 m)

Pressure at top of tower = 0.8 psia

Feed stage = 7

HVGO drawoff stage = 6

LVGO drawoff stage = 2

Off gas drawoff stage = 1

Hole diameter = 2 mm

Sieve hole area to active area = 0.12

Weir height = 0.4 in (1.0 cm)

Furnace duty = 692.3 Btu/hr = 174.5 kcal/hr.

Pump Around Result

I. Top pump around

Draw stage = 2

Return stage = 1

Temperature = 128^0F (53^0C)

Heat duty = -573.66 Btu/hr

Pressure = 0.8

Flow rate = 7.21 Ib/hr (3.27 kg/hr)

II. Bottom pump around

Draw stage = 6

Return stage = 5

Temperature = 420^0F (216^0C)

Heat duty = -482.59 Btu/hr

Pressure = 0.8

Flow rate = 5.89 Ib/hr (2.67 kg/hr)

CHAPTER FOUR

4.0 EQUIPMENT DESIGN

4.1 Actual Column Diameter for ADU Column and Strippers

Sketch indicating terms used in equipment design of column is shown in Figure 4.1.

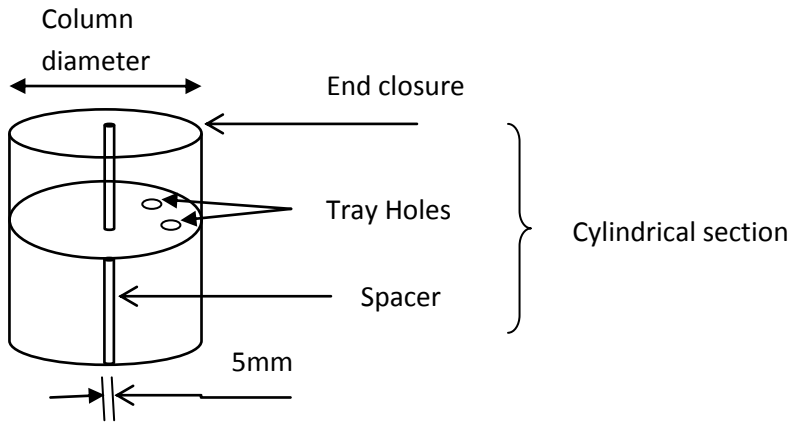


Figure 4.1: Sketch showing terms used in design

For distillation columns, the stacked type of construction is used where the column diameter is too small for a man to enter to assemble the plates, say less than 1.2 m (4 ft) (Sinnott 2006). Each plate is fabricated complete and then joined together by a rod called the spacer as shown in Figure 4.1.

Taking spacer diameter of 5 mm.

Cross Sectional Area of Spacer = $\frac{\pi d^2}{4}$ - - - - - - 4.1

$$= \frac{\pi(0.005)^2}{4}$$

$$= 0.0000196 \text{ m}^2$$

For ADU Column,

Where ADU Column Diameter = 7.5 cm from previous result (section 3.3.3).

Required ADU Column Area=$\frac{\pi(0.075)^2}{4}$

$= 0.00442\ m^2$

Two spacers are recommended (Sinnott 2006).

Actual ADU Column Area = Required Area + 2 x Area of Spacer - - 4.2

$= 0.00442 + 2 \times 0.0000196$

$= 0.00446\ m^2$

Actual Diameter of ADU Column= $\sqrt{\left(\frac{4 \times \text{Actual area}}{\pi}\right)}$ - - - - 4.3

$= \sqrt{\left(\frac{4 \times 0.00446}{\pi}\right)}$

$= 0.0754\ m$

For stripper,

Where, stripper Column Diameter = 2 cm from previous result (Section 3.3.3)

Required Stripper Column Area = $\frac{\pi(0.02)^2}{4}$

$= 0.000314\ m^2$

Actual Stripper Column Area = Required Area + 2 x Area of Spacer

$= 0.000314 + 2 \times 0.0000196$

$= 0.000353\ m^2$

Actual Diameter of Stripper Column= $\sqrt{\left(\frac{4 \text{ x Actual area}}{\pi}\right)}$

$= \sqrt{\left(\frac{4 \text{ x } 0.000353}{\pi}\right)}$

= 0.0212 m

4.2 Calculation of ADU Column Thickness

Material selected for use is stainless steel because of its high resistance to rust and great strength.

To calculate thickness of a column, parameters required are as follows:

Design temperature = temperature in column

Design pressure = internal pressure – external pressure of column

Typical design stress (f) which is a function of column material and design temperature. Typical design stress at various design temperature for stainless steel is shown in Table 4.1.

C = a Constant, which depends on the edge support. For diameters less than 0.6 m, flanged-only end is used with C = 0.45

Table 4.1: Typical Design Stresses for Stainless Steel 304 plate

Material	Tensile strength (N/mm^2)	Design stress at temperature ^0C (N/mm^2)							
		0 to 50	100	150	200	250	300	350	400
Stainless steel (304)	510	165	145	130	115	110	105	100	100

Source: Sinnott 2006

Design Temperature for ADU Column = 295^0C (highest temperature in column as shown in result presented in Table 3.33)

From Table 4.1 at 295^0C, Typical Design stress, f = 105 N/mm^2

Operating Pressure for ADU column = 2.758 bar

Design pressure, P taken as 10 per cent above operating pressure,

= (2.758 – 1) x 1.1

= 1.9338 bar

Design pressure for ADU column = 0.1934 N/mm^2

4.2.1 Cylindrical Section Thickness for ADU

Thickness, $E_{\text{ADU Cylindrical}} = \frac{PD}{4f-P}$ - - - - - 4.4

Sinnott, 2006.

Where D is Internal Diameter = 75.4 mm

$E_{ADU\ Cylindrical} = \frac{0.1934 \times 75.4}{4 \times 105 - 0.1934}$ = 0.035 mm

Add Corrosion allowance 0.035 + 1 = 1.2 mm (for safety reasons)

4.2.2 End Closures Thickness for ADU

Let a flat plate be used for closure,

Thickness, $E_{ADU\ End} = C \times D\sqrt{\frac{P}{f}}$ - - - - - - 4.5

Sinnott, 2005.

Where, diameter,D = 75.4 mm

Adopting the flanged-only end for the dome plate closure, C = 0.45

$E_{ADU\ End} = 0.45 \times 75.4\sqrt{\frac{0.1934}{105}}$ = 1.46 mm

Add Corrosion allowance 1.5 + 1 = 2.5mm.

4.3 Calculation ADU Stipper Thickness

Stripper design for stripper with highest temperature and pressure applies for all three stripper since their design parameters are within close range.

Material selected for use is stainless steel

Design Temperature = 275^0C

From Table 4.1 at 275^0C, typical design stress, f = 107.5 N/mm^2

Operating pressure for stripper = 2.57 bar

Design Pressure, P taken as 10 percent above operating pressure,

= (2.57 – 1) x 1.1

= 1.727 bar

P = 0.1727 N/mm^2

4.3.1 Cylindrical Section Thickness for Stripper

Thickness, $E_{\text{Stripper Cylindrical}} = \frac{PD}{4f-P}$

Where D is Internal Diameter = 20 mm

$E_{\text{Stripper Cylindrical}} = \frac{0.1727 \times 20}{4 \times 107.5 - 0.1727}$ = 0.008 mm

Add corrosion allowance 0.08 + 1 = 1.2 mm (for safety reasons)

4.3.2 End Closures Thickness for Stripper

Let a dome plate be used for closure,

Thickness, $E_{\text{Stripper End}} = C \times D\sqrt{\frac{P}{f}}$

Sinnott, 2006.

Adopting the flanged-only end for the dome plate closure, C = 0.45

$E_{\text{Stipper End}} = 0.45 \times 20\sqrt{\frac{0.1727}{107.5}}$ = 0.36 mm

Add Corrosion allowance 0.36 + 1 = 1.5 mm.

NOTE: Detail stress analysis is not carried in consideration of external loads such as insulator weight and factor such as wind but thickness were rounded up appropriately to make up for these factors.

4.4 Number of Holes for ADU Tray (Plate)

4.4.1 Number of Holes for ADU Main Column Tray

Sieve Hole Area = 0.12 x Active area - - - - - 4.6

(Aspen Plus simulation)

Downcomer Area/Column Area = 0.1 - - - - 4.7

Column Diameter = 0.0747 m

Column Area = $\boldsymbol{\pi}$ x $(0.0747^2)/4$ = 0.00438 m^2

Downcomer area = 0.000438 m^2

Active Area = Column area – 2 x Downcomer area - - - - 4.8

= 0.00438 – 2 x 0.000438

= 0.00351 m^2

Sieve Hole Area = 0.12 x 0.00351

= 0.0004208 m^2

Let Plate Hole Size = 2 mm diameter

One Hole Area = $\boldsymbol{\pi}$ x $(0.002^2)/4$ = 0.00000314 m^2

Number of Holes = $\frac{Sieve\ hole\ area}{one\ hole\ area}$ - - - - - - 4.9

Sinnott, 2005.

No of Holes for each ADU plate = $\frac{0.0004208}{0.00000314}$ = 134 holes

4.4.2 Number of Holes for All Three Stripper Tray

Sieve hole area = 0.12 x Active area (Aspen Plus simulation)

Downcomer area/column area = 0.1

Column Diameter = 0.02 m

Column Area =0.000314 m^2

Downcomer Area = 0.0000314 m^2

ACTIVE area = Column area – 2 x Downcomer area

Sinnott, 2005.

= 0.000314 – 2 x 0.0000314

= 0.000251 m^2

Sieve Hole area = 0.12 x 0.000251

= 0.00003012 m^2

Let Plate hole size = 2 mm diameter

One Hole area = $\boldsymbol{\pi}$ x $(0.002^2)/4$ = 0.00000314 m^2

Number of Holes = $\frac{Sieve\ hole\ area}{one\ hole\ area}$

No of Holes for each stripper tray = $\frac{0.00003012}{0.00000314}$ = 9.6 ≈ 10 holes

4.5 Pipe Diameter Specification

For carbon steel pipe

Optimum pipe diameter (d, opt) = $293\ G^{0.53}\ \ell^{-0.37}$ - - - - 4.10

Where G = Mass flow rate

ℓ = density of fluid

For crude flow line

G = 5.687 kg/hr

ℓ = 860 kg/m^3

d, opt = $293 \times \left(\frac{5.687}{3600}\right)^{0.53} 860^{-0.37}$

Pipe diameter for crude flow line, d, opt = 0.79 mm ≈1 mm

This is not a standard size pipe, closest that can be possible found is 5 mm.

4.6 Furnace Design for ADU

Box still pipe furnace was selected because simplicity in design, fabrication and operation. A sketch of box still pipe is shown in Figure 4.2.

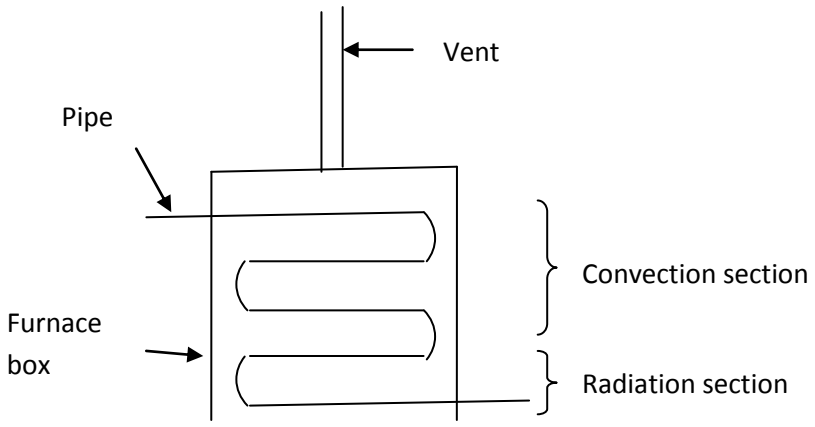


Figure 4.2: sketch of a box still pipe furnace

Heat duty of ADU furnace = 981.4 kcal/hr. (obtained from ASPEN Plus simulation design result)

Total heat with 18% losses = 981.4 + (0.18 x 981.4)

= 1158.05 kcal/hr

Heat value of fuel (Natural gas) = 12 000kcal/kg (Rao,1990)

Amount of fuel required for ADU furnace $= \frac{Total\ heat}{Heat\ value} = \frac{1158.05}{12000}$

= 0.0965 kg/hr

The furnace consists of convection and radiation section. Calculations are carried out for sections separately.

4.6.1 Convective Section for ADU Furnace

Let 30% heat be absorbed in convection section.

Outlet Temperature of crude in convection section = (350 – 75) x (0.3)

= 82.5

i.e. 48 + 75 = 157.5^0C

Amount of fuel = 0.0965 kg/hr

For excess air available, air/gas ratio = 25 - - - - - 4.11

Rao, 1990.

Amount of air used = 0.0965 x 25 = 2.41 kg/hr

Total gas = Air used + Fuel - - - - - - - 4.12

=2.41 + 0.0965 kg/hr

= 2.51 kg/hr

For gas

$Qc = mC_p\Delta t$ - - - - - - - - 4.13

Rao, 1990.

1cal = 4.184J

Heat duty, Q =1158.05 kcal/hr x 4.184 = 4 845.28 kJ/hr

0.3 x 4 845.28 = 2.51 x 1.374 x Δt

$\Delta t = 421.5^0C$

Gas inlet temperature t = 421.5 + 30 = 451.5^0C

From countercurrent flow temperature curve

$\Delta t1 = 451.5 – 157.5 = 294^0C$

$\Delta t2 = 200^0C$ (Rao, 1990)

LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$ - - - - - - - - 4.14

= 244^0C

Ac = $\frac{Qc}{U(LMTD)}$ - - - - - - - - - 4.15

= $\frac{0.3 \times 4845.28}{85.86 \times 244}$

= 0.0694 m^2

Let length of tube, L = 0.4m

No of tubes = $\frac{A}{\pi DL} = \frac{0.0667}{\pi \times 0.005 \times 0.4}$

No of tubes for ADU furnace convection section = 12 tubes

4.6.2 Radiation Section for ADU Furnace

For Crude

$mC_p\Delta t = Q_R$ Rao, 1990.

Mass of crude, m = 5.6865 kg/hr

Specific heat capacity, C_p = 2.268 kJ/kg^{0}C

Q_R = Heat absorped by Radiation section

5.6865 x 2.268 x (350 – 157.5) = R x 4845.28

R = 0.51

$R = \frac{1}{1+G\frac{\sqrt{\frac{Q}{\propto A}}}{S}}$ - - - - - - - - - 4.16

Rao, 1990.

Where constants s = 14200, $\propto$ = 0.88

$0.51 = \frac{1}{1+25\frac{\sqrt{\frac{4845.28}{\propto A}}}{14200}}$

$\propto$A = 0.0163

A = N x D x space x L - - - - - - - 4.17

= N x 0.005 x 2 x 0.4 = N x 0.004 (space = 2D)

0.0163 = 0.88 x N x 0.004

Number of tubes for ADU radiation section, N = 5 tubes

4.7 Pump Design for ADU

Velocity of crude, u = $\frac{1 \times 159 \times 4}{3600 \times 24 \times \pi \times 1000 \times (0.005)^2}$ = 0.1 m/s

Reynolds Number = $\frac{\rho du}{\mu}$ - - - - - - 4.18

Where

ρ = Density

d = Diameter

u= Velocity

μ = Viscosity

Reynolds Number= $\frac{0.1 \times 0.005 \times 860}{0.01 \times 10^{-4}}$

= 4.3 x 10^5

Reynolds Number shows that flow is in turbulent region.

$P_C + P_R = \frac{4fL\ell V^2}{D}$ - - - - - - - - 4.19

Sinnott, 2005.

From chart, f = Fanning friction factor = 0.002

$P_C + P_R = \Delta P_f$ - - - - - - - - - 4.20

$= \frac{4\times0.002\times0.4\times(12+4)\times(0.1)^2}{0.005} = 0.1024\ N/m^2$

Taking total energy balance for crude storage, pump and furnace

$g\Delta z + \frac{\Delta P}{\rho} - \frac{\Delta P_f}{\rho} - W = 0$ - - - - - - - 4.21

Sinnott, 2005.

where Δz is difference in elevation,m

ΔP is difference in system pressure,N/m^2

W is Work done by pump.

$g\Delta z$ and $\frac{\Delta P_f}{\rho}$ are negligible.

$-W = -\frac{\Delta P}{\rho}$ - - - - - - - - - 4.22

Sinnott, 2005.

$-W = -\frac{(103421.36 - 166715.232)}{860}$

$= 77.2$ J/kg,

$\text{Power} = \frac{(W \times m)}{\eta}$ - - - - - - - - 4.23

M = mass flow rate = 5.687 kg/hr

Let η = efficiency = 0.7

$\text{Power for pump} = \frac{(77.2 \times 5.687)}{0.7}$

= 627.195 W = 0.63 kW

Power for ADU pump = 0.845 hp ≈ 1 hp.

4.8 Condenser Design for ADU

From previous result, Condenser duty = Q = 3426.82 Btu/hr = 1004.3 W = 1 kW

For cooling water flow rate

$Q = mc\Delta T$

Sinnott, 2005.

C = Specific heat capacity of water = 4.18 kJ/kg^{0}C

Let cooling water be available at 30^{0}C and the temperature rise is to be limited to 10^{0}C.

Cooling water flow = $\frac{1}{(40-30)4.18}$ = 0.024 kg/s

From Table 12.1 (Sinnott 2005),

Assume overall coefficient = 900 W/m^{20}C

We design for a horizontal shell and tube heat exchanger.

$A = \frac{Q}{U(LMTD)}$

Sinnott, 2005.

Log. Mean Temperature Difference, LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log\frac{\Delta t1}{\Delta t2}}$

$\Delta t1 = 117 - 40 = 77$

$\Delta t2 = 53 - 30 = 23$

LMTD = $\frac{77 - 23}{2.303 log\frac{77}{23}}$

$= 44.7^0C$

$A = \frac{1004.3}{900(44.7)} = 0.025\ m^2$

Surface area of one tube = **π**DL

Sinnott, 2005.

Take D = 6mm o.d.

L = 0.5 m

Surface area of one tube = **π** x 0.006 x 0.5 = 0.00942 m^2

Number of tubes in condenser = $\frac{0.025}{0.00942}$ = 2.65 ≈3 tubes

4.8.1 Shell Inside Diameter for Condenser

Since number of tube is three, we use triangular tube arrangement.

Most recommended tube pitch (distance between tube centres) is 1.25 times the tube outside diameter, OD. Where OD = 6 mm (Sinnott, 2006)

Bundle diameter = $D_b = OD\left(\frac{N}{K}\right)^{\frac{1}{n}}$ - - - - - - 4.24

Sinnott, 2006.

N = 3,

Constants K = 0.319 and n = 2.142 (Sinnott, 2006)

$D_b = 6\left(\frac{3}{0.319}\right)^{\frac{1}{2.142}}$ = 17 mm = 0.

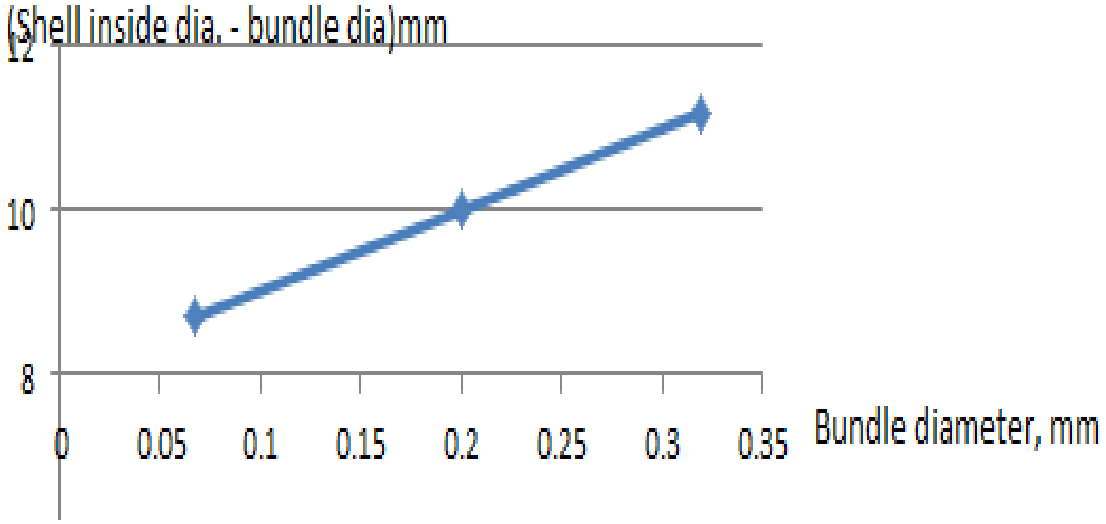


Figure 4.1: Plot of (shell inside diameter – bundle diameter)(mm) vs Bundle diameter (m). Source: Sinnott 2006.

From Figure 4.1, at bundle diameter = 0.017 m, shell inside diameter – bundle diameter ≈ 8 mm

Therefore, shell inside diameter = 8 + 17 = 25 mm

4.9 Crude Preheater Heat Exchanger for ADU
1 kW = 3412.14 Btu/hr

Heat exchanger duty = Q = 482.59 Btu/hr = 0.1416 kW

Required Exchange Area = 0.00188 m^2

Surface area of one tube = **π**DL

Sinnott, 2006.

Tube diameter, D = 6 mm outer diameter

L = 0.3 m

Surface area of one tube = **π** x 0.006 x 0.3 = 0.00565 m^2

Number of tubes = $\frac{0.00188}{0.00565}$ = 0.33 ≈1 tube

4.9.1 Shell Inside Diameter for Crude Preheater Heat Exchanger

Since number of tube is one, bundle diameter = tube outer diameter = 0.006 m

From Figure 4.1, shell inside diameter – bundle diameter = 8 mm

Therefore, shell inside diameter = 8 + 6 = 14 mm

4.10 Design of Reflux Drum for ADU

The reflux drum is a three phase separator, separating gas, light naphtha and water, since gas constituent is 1% of crude by volume, gas separation is negliable. Reflux drum is taken to be liquid- liquid separator (Onwuka, 2010). Relevant equations are as follows:

Derived from Stokes' law, capacity of separator is given below:

Vertical Vessels:

$$W_{cl} = C\left(\frac{Sh-Sl}{\mu}\right)(0.785)D_v^2 \quad - - - - - - - \quad 4.25$$

Horizontal Vessels:

$$W_{cl} = C\left(\frac{Sh-Sl}{\mu}\right)L_l H_l \quad - - - - - - - - \quad 4.26$$

Separator volume is given as:

$$V = \frac{W(t)}{1440} \quad - - - - - - - - - \quad 4.27$$

Onwuka, 2010

W_{cl} = Flow rate of hydrocarbon

Sh = Specific gravity of water

Sl = Specific gravity of hydrocarbon

$\boldsymbol{\mu}$ = Viscosity of hydrocarbon

Dv = Diameter of separator

L_l = Length of separator

H_l = Height of separator

V = Volume of separator

W = Total flow rate

t = Retention time

C = Constant given in Table 4.1 which shows level of emulsion in liquid mixture.

Table 4.1: Values of C

Emulsion Characteristic	Droplet Diameter, Microns	Constant C
Free liquids	200	1100
Loose emulsion	150	619
Moderate emulsion	100	275
Tight emulsion	60	99

Source: Onwuka, 2010.

4.10.1 Sizing of Reflux Drum

Light naphtha flow rate = reflux rate + distillate rate - - - 4.28

= 0.85688987 + 0.2051189 = 1.062 bbl/day (7.036 litre/hr).

Free water rate = 0.0763 bbl/day (0.505 litre/hr)

Operating temperature = 127^0F (52.8^0C)

Operating pressure = 15 psia

Water specific gravity = 1.0

Light Naphtha specific gravity = 0.7525

Crude oil viscosity =0.55 cp

Retention time = 5 min

C = 1100 (water exist as free liquid)

By Equation 4.25

$1.1 = 1100\frac{1-0.7525}{.55}(0.785)(D_v)^2$

$D_v = 0.053\text{ft} \approx 1\text{ ft} = .03\text{ m}$

Lets take Reflux drum diameter, D to be 0.05m

By Equation 4.27

$V = \frac{1.14 \times 5}{1440} = 0.00396\text{bbl} = 0.629\text{ dm}^3$

$= 0.000629\text{ m}^3$

$V = \frac{H\pi D^2}{4}$ (Volume of cylinder)

$0.000629 = \frac{H\pi 0.05^2}{4}$

Height of Reflux drum, H = 0.32 m

Size of reflux drum diameter x height is 5cm x 32cm respectively.

4.10.2 Calculation of Reflux Drum Thickness

Material selected for use is stainless steel for its high corrosion resistance and strength.

Design temperature ≈ 50^0C (from Aspen Plus simulation result)

From (Sinnott 2005), Typical design stress, f = 165 N/mm^2

Operating pressure of reflux drum is at atmospheric pressure = 1.034 bar.

Design pressure; taken as 10 percent above operating pressure,

= (1.034 – 1) x 1.1

= 0.0374 bar

Design pressure of reflux drum = 0.00374 N/mm^2

Cylindrical Section Thickness for Reflux drum

Thickness, $E_{\text{Reflux drum cylindrical}} = \frac{PD}{4f-P}$

Sinnott, 2005.

Where D is internal diameter = 50 mm

$E_{\text{Reflux drum cylindrical}} = \frac{0.00374 \times 50}{4 \times 165 - 0.00374}$ = 0.00028 mm

Add corrosion allowance 0.00028 + 1 = 1.2 mm (for safety, next standard after 1 mm is selected)

End Closures Thickness for Reflux drum

Let a dome plate be used for closure,

Thickness, $E_{\text{Reflux drum end}} = C \times D\sqrt{\frac{P}{f}}$ Sinnott, 2005.

Adopting the flanged-only end for the dome plate closure, C = 0.45

$E_{\text{Reflux drum end}} = 0.45 \times 50\sqrt{\frac{0.00374}{165}} = 0.11$ mm

Add corrosion allowance 0.11 + 1 = 1.11 = 1.2 mm (closest standard size).

4.11 Calculation of Actual Column Diameter for VDU
Let spacer diameter be 5 mm.

Cross sectional area of spacer = $\frac{\pi d^2}{4}$

$$=\frac{\pi(0.005)^2}{4}$$

$$= 0.0000196 \text{ m}^2$$

For main column,

Required area = $\frac{\pi(0.099)^2}{4}$

$$= 0.00769 \text{ m}^2$$

Actual Area = required area + 2 x area of spacer

= 0.00769 + 2 x 0.0000196

= 0.00773 m^2

Actual diameter of VDU column = $\sqrt{\left(\frac{4 \times Actual\ area}{\pi}\right)}$

= $\sqrt{\left(\frac{4 \times 0.00773}{\pi}\right)}$

= 0.0992 m

4.12 Calculation of VDU Main Column Thickness

Material selected for use is stainless steel because of high corrosion resistance and strength.

Design temperature = 332^0C

Young's modulus of steel, E = 2 x 10^{11} N/mm^2

Design pressure, taken as 10 percent above external pressure,

= 1 x 1.1

= 1.1 bar

P = 0.11 N/mm^2

Thickness of VDU Column, e = $4R\sqrt{\frac{P}{E}}$

Sinnott, 2005.

= 4 X 49.5$\sqrt{\frac{.11}{200000}}$

= 0.15 mm

Add corrosion allowance 0.15 + 1 = 1.5 mm.

4.13 Number of Holes Per Tray for VDU

Sieve hole area = 0.12 x Active area (Aspen Plus simulation)

$\frac{\text{Downcomer area}}{\text{column area}} = 0.1$

Column area = 0.00769 m^2

Downcomer area = 0.000769 m^2

Active area = Column area – 2 x downcomer area

= 0.00769 – 2 x 0.000769

= 0.006152 m^2

Sieve hole area = 0.12 x 0.006152

= 0.000738 m^2

Let plate hole size = 2 mm diameter

One hole area = $\frac{\pi(0.002)^2}{4}$

= 0.00000314 m^2

Number of holes = $\frac{Sieve\ hole\ area}{one\ hole\ area}$

No of holes for VDU tray = $\frac{0.000738}{0.00000314}$ = 236 holes

4.14 Furnace Design for VDU

Heat duty of VDU furnace = 174.5 kcal/hr (obtained from ASPEN Plus simulation design result)

About 18% of heat is lost in a standard furnace (Rao, 1990). So 18% of heat is added to make up for the loss.

Total heat with 18% losses = 174.5 + (0.18 x 174.5)

= 205.91 kcal/hr

Heat value of fuel (Natural gas) = 12 000kcal/kg (Rao, 1990)

Amount of fuel required for VDU furnace = $\frac{Total\ heat}{Heat\ value} = \frac{205.91}{1200}$

= 0.0172 kg/hr

The furnace consists of convection and radiation section. Calculations are carried out for sections separately.

4.14.1 Convective Section for VDU

Let 30% heat be absorbed in convection section.

Outlet temperature of crude in convection section = (373 – 294) x (0.3)

= 23.7

i.e. 23.7 + 294 = 317.7^0C

Amount of fuel = 0.0172 kg/hr

For excess air available, air/gas ratio = 25

Amount of air used = 0.0172 x 25 = 0.43 kg/hr

Total gas = air used + fuel = 0.43 + 0.0172 kg/hr

= 0.45 kg/hr

For gas

$Qc = mC_p\Delta t$

Mass flow rate,m = 0.45kg/hr (Total gas calculated above)

Specific Heat Capacity of gas, C_p = 1.374 KJ/kg.K

1cal = 4.184J (conversion factor)

Heat duty, Q =205.91 kcal/hr x 4.184 = 861.53 kJ/hr

0.3 x 861.53 = 0.45 x 1.374 x Δt

$\Delta t = 418^0C$

Gas inlet temperature t = 418 + 30 = 448^0C

From countercurrent flow temperature curve

$\Delta t1 = 448 – 317.7 = 130.3^0C$

$\Delta t2 = 200^0C$ (Rao 1990)

Log Mean Temperature Difference, LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$

$= 163^0C$

$Ac = \frac{Qc}{U(LMTD)} = \frac{0.3 \times 861.53}{85.86 \times 163}$

$= 0.0185\ m^2$

Let Length of tube, L = 0.4m

Diameter of tube, D = 5 mm

Number of tubes for convective section of VDU furnace = $\frac{A}{\pi DL} = \frac{0.0185}{\pi \times 0.005 \times 0.4}$

= 3 tubes

4.14.2 Radiation Section for VDU

Since 30% heat is absorbed in convective section and 18% heat loss, then 52% is absorbed in the radiation section.

$$R = \frac{1}{1+G\frac{\sqrt{\frac{Q}{\propto A}}}{S}}$$

Where constants s = 14200, $\propto$ = 0.88

$$0.52 = \frac{1}{1+25\frac{\sqrt{\frac{861.53}{\propto A}}}{14200}}$$

$\propto$A = 0.003134

A = N x D x space x L = N x 0.005 x 2 x 0.4 = N x 0.004 (space = 2D)

0.003134 = 0.88 x N x 0.004

Number of tube in VDU radiation section, N = 0.89 = 1 tube

4.15 Design of Pump around Heat Exchanger for VDU

Design was done for TPA and BPA Heat exchangers as shown below:

4.15.1 TPA Heat Exchanger

1 kW = 3412.14 Btu/hr (conversion factor)

Heat exchanger duty = Q = 573.66 Btu/hr = 0.168 kW (from ASPEN Plus simulation result)

For cooling water flow rate

$Q = mc\Delta T$ Sinnott, 2006.

Where,

m = Mass flow rate

ΔT = Temperature difference

C = Specific heat capacity of water = 4.18 kJ/kg^{0}C

Let cooling water be available at 30^{0}C and the temperature rise is to be limited to 30^{0}C because of environmental standards.

Cooling water flow = $\frac{0.168}{(60-30)4.18}$ = 0.00134 kg/s

From Table 12.1 in Sinnott 2005,

Take overall coefficient = 900 W/m^{20}C

We design for a horizontal shell and tube heat exchanger.

Heat exchange Area, A = $\frac{Q}{U(LMTD)}$

Log Mean Temperature Difference, LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$

Sinnott, 2005.

$\Delta t1 = 140.6 - 60$

$= 80.6^0$C

$\Delta t2 = 53.3 - 30$

$= 23.3^0C$

LMTD for TPA Heat exchanger = $\frac{80.6 - 23.3}{2.303 log \frac{80.6}{23.3}}$

$= 71.1\ ^0C$

Heat exchange area for TPA Hx, A = $\frac{168}{900(71.1)}$

$= 0.00262\ m^2$

Surface area of one tube = **π**DL Sinnott, 2005.

Where,

D = Diameter of pipe

L = Length of pipe

Take D = 6 mm o.d.

L = 0.2 m

Surface area of one tube = **π** x 0.006 x 0.2

$= 0.00377\ m^2$

Number of tubes for TPA Heat exchanger = $\frac{A}{Surface\ area\ of\ one\ tube} = \frac{0.00262}{0.00377}$

$= 0.7 \approx 1$ tube

4.15.2 BPA Heat Exchanger

Heat exchanger duty = Q = 482.59 Btu/hr = 0.141 kW (from ASPEN Plus simulation result)

For cooling water flow rate

$Q = mc\Delta T$

C = specific heat capacity of water = 4.18 kJ/kg^{0}C

Let cooling water be available at 30^{0}C and the temperature rise is to be limited to 30^{0}C because of environmental standards.

Cooling water flow = $\frac{0.141}{(60-30)4.18}$ = 0.00112 kg/s

From Table 12.1 in Sinnott 2005,

Take overall coefficient = 900 W/m^{20}C

We design for a horizontal shell and tube heat exchanger.

$A = \frac{Q}{U(LMTD)}$ Sinnott, 2005.

$LMTD = \frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$ Sinnott, 2005.

$\Delta t1 = 287 - 60$

$= 227^0C$

$\Delta t2 = 216 - 30$

$= 186^0C$

LMTD for BPA Heat exchanger = $\frac{227-186}{2.303log\frac{227}{186}}$

= 205.8 ^{0}C

Heat exchange area for BPA Hx, A = $\frac{141}{900(205.8)}$

= 0.000761 m^2

Surface area of one tube = **π**DL

Take Diameter, D = 6mm o.d.

Length, L = 0.2 m

Surface area of one tube = **π** x 0.006 x 0. 2

= 0.00377 m^2

Number of tubes for BPA Heat exchanger= $\frac{A}{Surface\ area\ of\ one\ tube} = \frac{0.000761}{0.00377}$

= 0.2 ≈1 tube

4.15.3 Shell Inside Diameter for TPA Heat Exchanger

Since number of tube is one, bundle diameter = tube diameter = 0.005 m

From figure 4.1, shell inside diameter – bundle diameter = 8 mm

Therefore, shell inside diameter for TPA Heat exchanger = 8 + 5

= 13 mm

4.15.4 Shell Inside Diameter for BPA Heat Exchanger

Since number of tube is one, bundle diameter = tube diameter = 0.005 m

From plot, shell inside diameter – bundle diameter = 8 mm

Therefore, shell inside diameter for BPA Heat exchanger = 8 + 5

= 13 mm

4.16 Summary of Equipment Design Result

Summary of furnaces, columns, condenser and heat exchanger are presented below:

4.16.1 Summary of Furnace Design Result

Furnace Data Sheet

Mechanical data for ADU and VDU furnaces is shown in Table 4.2.

Table 4.2 Mechanical Data for ADU and VDU Furnaces

	ADU Furnace		VDU Furnace	
Furnace type	Box pipe still heater		Box pipe still heater	
	Design	Actual	Design	Actual
No of tubes in convection section	12	12	-	-
No of rows in convection section	4	3	-	-
No of tubes in radiation section	5	5	4	5
No of rows in radiation section	1	1	1	1
Diameter of tube, mm	5	13	5	13
Length of tube, m	0.4	0.4	0.4	0.4
Spacing between tubes	2D	+2D	2D	+2D
Material of construction	Stainless steel	Mild steel	Stainless steel	Mild steel
Insulation thickness, cm	7.5	5	7.5	5

Technical/Operational Data for ADU and VDU Furnaces is shown in Table 4.3.

Table 4.3: Technical/Operational Data for Furnaces

S/N		ADU Furnace	VDU Furnace
1	Heat duty, Btu/hr	3894	695
2	Temperature, ^{0}F	608	703
	^{0}C	320	373
3	Pressure, Psi	24	2.03
4	Feed flow, ft^3/hr	0.2338	0.0816
	m^3/hr	0.0066	0.0023
5	Vapour flow, ft^3/hr	0.1298	0.0483
	m^3/hr	0.0037	0.0014
6	Liquid flow, ft^3/hr	0.1040	0.0338
	m^3/hr	0.0029	0.0010
7	Fluid material	Crude oil	Atmospheric residue
8	Percentage heat loss, %	18	18
9	Fuel heat value, kcal/kg	12000	12000
10	Fuel flow rate, kg/hr	0.0965	0.0172
11	Outlet temperature of crude in convection section, ^{0}C	157.5	317.7
12	Log Mean Temp. Difference (LMTD), ^{0}C	244	163

4.16.2 Summary of Column Design Result

Columns Data Sheet

Mechanical Data for Strippers, ADU and VDU Column is shown in Table 4.4.

Table 4.4: Mechanical Data for Strippers, ADU and VDU Column

S/N	Parameter	Stripper 1, 2 & 3		ADU Column		VDU Column	
		Design	Actual	Design	Actual	Design	Actual
1	Tray type	Sieve	sieve	sieve	Sieve	sieve	sieve
2	Number of pass	1	1	1	1	1	1
3	Number of spacer	2	1	2	1	2	1
4	Spacer diameter, mm	5	8.5	5	8.5	5	8.5
5	Number of trays	4	4	20	20	9	9
6	Number of holes/tray	10	-	134	-	236	-
7	Column diameter, cm	2.1	2.5	7.5	9	9.9	9
8	Hole diameter, mm	2	2	2	2	2	2
9	Weir height, inch	1.5	No weir	1.5	1	0.4	1
	Cm	3.81	No weir	3.81	2.54	1.02	2.54
10	Weir length, cm	1.3	±1	5.4	±6	7.2	±6
11	Tray spacing, cm	15	15	15	15	15	15
12	Plate deflection, mm	< 3	< 3	< 3	< 3	< 3	< 3
13	Downcomer area, m^2	0.000031 4	-	0.0004 38	-	0.0007 69	-
14	Material of construction	Stainless steel	Mild steel	Stainles s steel	Mild steel	Stainles s steel	Mild steel
15	Column wall thickness, mm	1.2	3	1.2	3	1.5	3
16	Column cover thickness, mm	1.5	3	2.5	3	1.5	3
17	Insulation thickness, cm	7.5	10	7.5	10	7.5	10
18	Insulation material	Mineral Wool					

Technical/Operational Data for ADU column and VDU column are shown in Table 4.5 and Table 4.6 respectively.

Table 4.5: Technical/Operational Data ADU Column

S/N	Stream	Product/ Feed	Stage/Tray Number
1	Light Naphtha from ADU	Product	1
2	Heavy Naphtha from ADU	Product	6
3	kerosene from ADU	Product	13
4	Crude from furnace	Feed	18
5	AGO from ADU	Product	17
6	Atmospheric residue	Product	20
7	Steam to ADU	Feed	20

Table 4.6: Technical/Operational Data VDU Column

S/N	Stream	Product/ Feed	Stage Number
1	Off gas	Product	1
4	LVGO	Feed	2
5	HVGO	Product	6
6	Vacuum residue	Product	9
7	Steam to VDU	Feed	9
	Atmospheric residue from ADU	Feed	7

NOTE: All strippers have their respective crude product feed to Stage 3 and steam to Stage 4. Stripper leaves at the bottom of stripper distillate leaves at the top.

All stream properties are given in Process Flow Diagram data sheet.

Summary of Process Flow Result

Process stream flow data is presented in Table 4.7.

Table 4.7: Process Flow Data

S/N	Stream	Liq/vapour	Pressure, psi	Temperature,		Flowrate,	
				^{0}F	^{0}C	ft^3/hr	m^3/hr
1	Desalted crude	L	15.0	86	30	0.2363	0.00669
2	Light Naphtha	L	15.0	127	52.8	0.0504	0.00143
3	Heavy Naphtha	L	30.0	277	136.1	0.0276	0.00078
4	Kerosene	L	34.7	372	188.9	0.0335	0.00095
5	Cold AGO	L	37.3	150	65.6	0.0480	0.00136
6	AGO	L	37.3	488	253.3	0.0573	0.00162
7	Atmospheric residue	L	40.0	561.1	293.9	0.1076	0.00305
8	Hot crude	L	15.0	167	75	0.2463	0.00697
9	LVGO	L	0.7734	285	140.6	0.0425	0.00120
10	HVGO	L	0.7734	549	287.2	0.0364	0.00103
11	TPA reflux	L	0.7734	128	53.3	0.1289	0.00365
12	BPA reflux	L	0.7734	420	215.6	0.1009	0.00286
13	Steam to ADU	V	64.7	450	232.2	5.4826	0.15516
14	Steam to VDU	V	60.0	450	232.2	5.81	0.16442
15	Steam to S1	V	64.7	450	232.2	1.0224	0.02893
16	Steam to S2	V	64.7	450	232.2	1.2892	0.03648
17	Steam to S3	V	64.7	450	232.2	1.4373	0.04068
18	Gas	V	Cs	cs		cs	
19	Off gas to jet	V	0.7734	150	65.6	298.15	8.43765
20	Water	L	15.0	86	30	0.04726	0.00134
21	Waste water from DS	L	15.0	86	30	0.04726	0.00134
22	Waste water from RD	L	15.0	127	52.8	0.018	0.00051
23	Vacuum Residue	L	0.7734	555	290.6	0.0155	0.00044

*cs – controller specified.

4.16.3 Summary of Condenser/Heat Exchanger Design Result

Condenser/heat exchanger data sheet and their operating/ technical data are presented below:

4.16.3.1 Condenser/Heat Exchanger Data Sheet

Mechanical Data for Condenser and Heat exchangers is presented in Table 4.8.

Table 4.8: Mechanical Data Condenser/Heat exchanger (HX)

S/N		Condenser		Crude HX		TPA HX		BPA HX	
1	Type	Fixed-tube plate shell and tube heat exchanger for all four.							
		Design	Actual	Design	Actual	Design	Actual	Design	Actual
2	No of tube	3	3	1	1	1	1	1	1
3	Outer Diameter of tube,mm	6	13	6	13	6	13	6	13
4	Length of tube, m	0.5	0.5	0.3	0.3	0.2	0.2	0.2	0.2
5	Tube pitch	1.25	-	-	-	-	-	-	-
6	Tube pattern	Triangular	Triangular	-	-	-	-	-	-
7	Number of shells	1	1	1	1	1	1	1	1
8	Bundle diameter, mm	17	-	6	-	6	-	6	-
9	Inside Diameter of shell, mm	25	100	14	50	14	50	14	50
10	No of baffle	1	0	1	0	1	0	1	0
11	Baffle cut,mm	5	-	3	-	3	-	3	-
12	Material of construction				Design;Stainless steel for all four Actual; mild steel				

4.16.3.2 Technical/Operational Data for Condenser

i. Condenser Technical/Operational Data

Heat duty = 3426.82 Btu/hr

Exchange area = 0.025 m^2

LMTD = 44.7^0C

Avg. Heat transfer coefficient = 900 W/m^{20}C

Condenser Operational Data is shown in Table 4.9.

Table 4.9: Condenser Operational Data

	Shell side	Tube side
Fluid material	Water	Light naphtha + steam
Flow rate, kg/s	0.024	0.001575
Inlet temperature, 0C	30	117
Outlet temperature, 0C	40	53

ii. Crude Heat Exchanger Technical/Operational Data

Heat duty = 483.3 Btu/hr

Exchange area = 0.00188 m^2

LMTD = 88.6^0C

Average Heat transfer coefficient = 850 Watt/m^2K

Crude Heat Exchanger Operational Data is shown in Table 4.10

Table 4.10: Crude Heat Exchanger Operational Data

	Shell side	Tube side
Fluid material	Crude	AGO
Flow rate, bbl/day	1	0.3487
Inlet temperature, ^{0}C	30	273
Outlet temperature, ^{0}C	75	66

iii. TPA Heat Exchanger Technical/Operational Data

Heat duty = 3412.14 Btu/hr

Exchange area = 0.00262 m^2

LMTD = 71.1

Average Heat transfer coefficient = 900 W/m^{20}C

TPA Heat Exchanger Operational Data is shown in Table 4.11.

Table 4.11: TPA Heat Exchanger Operational Data

	Shell side	Tube side
Fluid material	Water	LVGO
Flow rate, kg/s	0.00134	0.000908
Inlet temperature, ^{0}C	30	140.6
Outlet temperature, ^{0}C	60	53.3

iv. BPA Heat Exchanger Technical/Operational Data

Heat duty = 3412.14 Btu/hr

Exchange area =0.000761 m^2

LMTD = 205.8^0C

Avg. Heat transfer coefficient = 900 W/m^{20}C

BPA Heat Exchanger Operational Data is shown in Table 4.12.

Table 4.12: BPA Heat Exchanger Operational Data

	Shell side	Tube side
Fluid material	Water	HVGO
Flow rate,kg/s	0.00112	0.000742
Inlet temperature	30	287
Outlet temperature	60	216

CHAPTER FIVE

5.0 DRAWINGS AND FABRICATION

5.1 Working Drawing for Mini Refinery Components

Working drawing was done for all components using AutoCAD Software. Both isometric and angle projection drawing were done for working drawing. Detail dimensioning was done on all drawings. All units are in mm and drawings are not to scale. Working drawings are shown in Figure 5.1 to Figure 5.6. Enlarge copies of working drawings on A3 size paper is attached to this work for clarity.

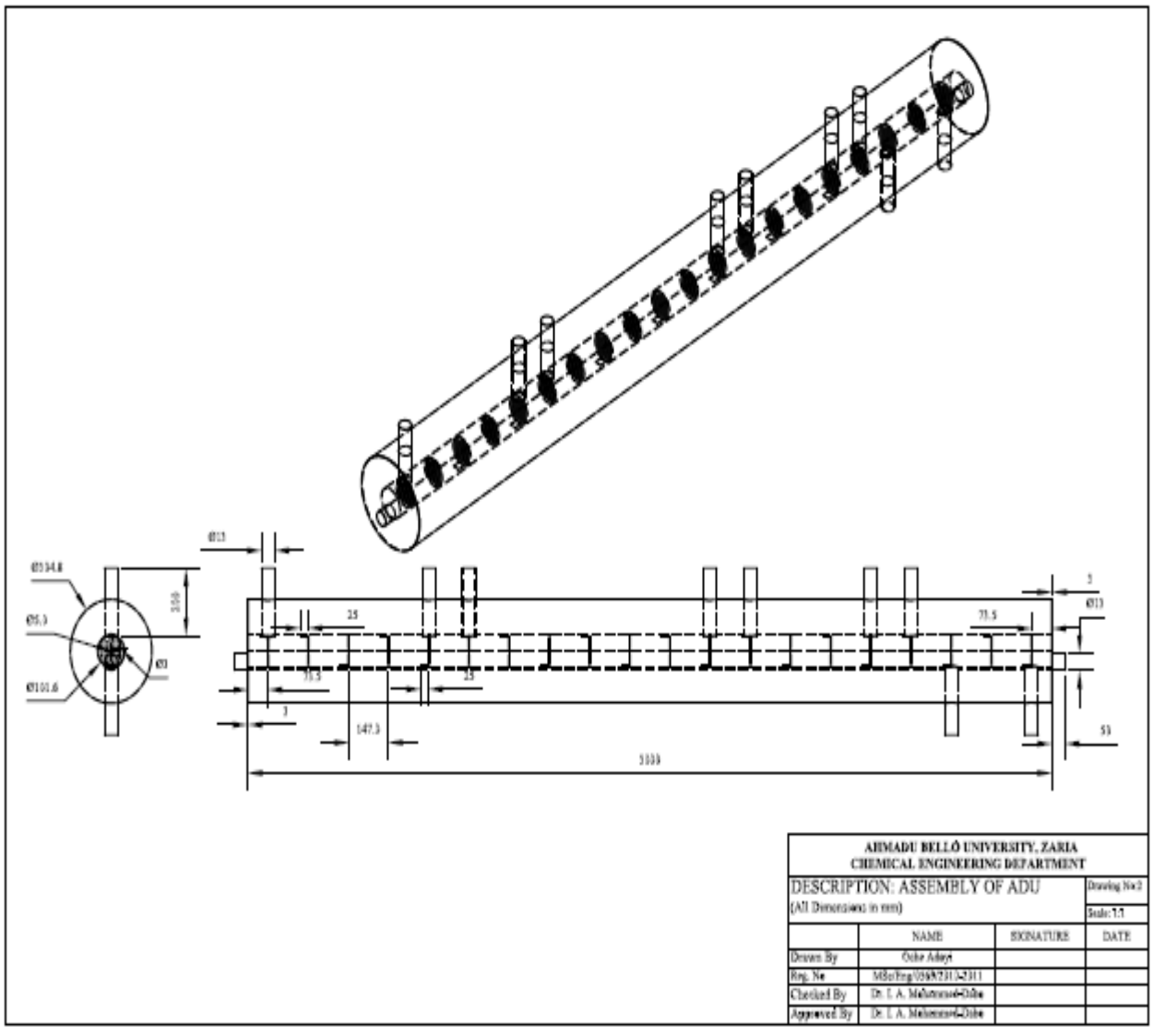


Figure 5.1 : Atmospheric Distillation Unit Working Drawing

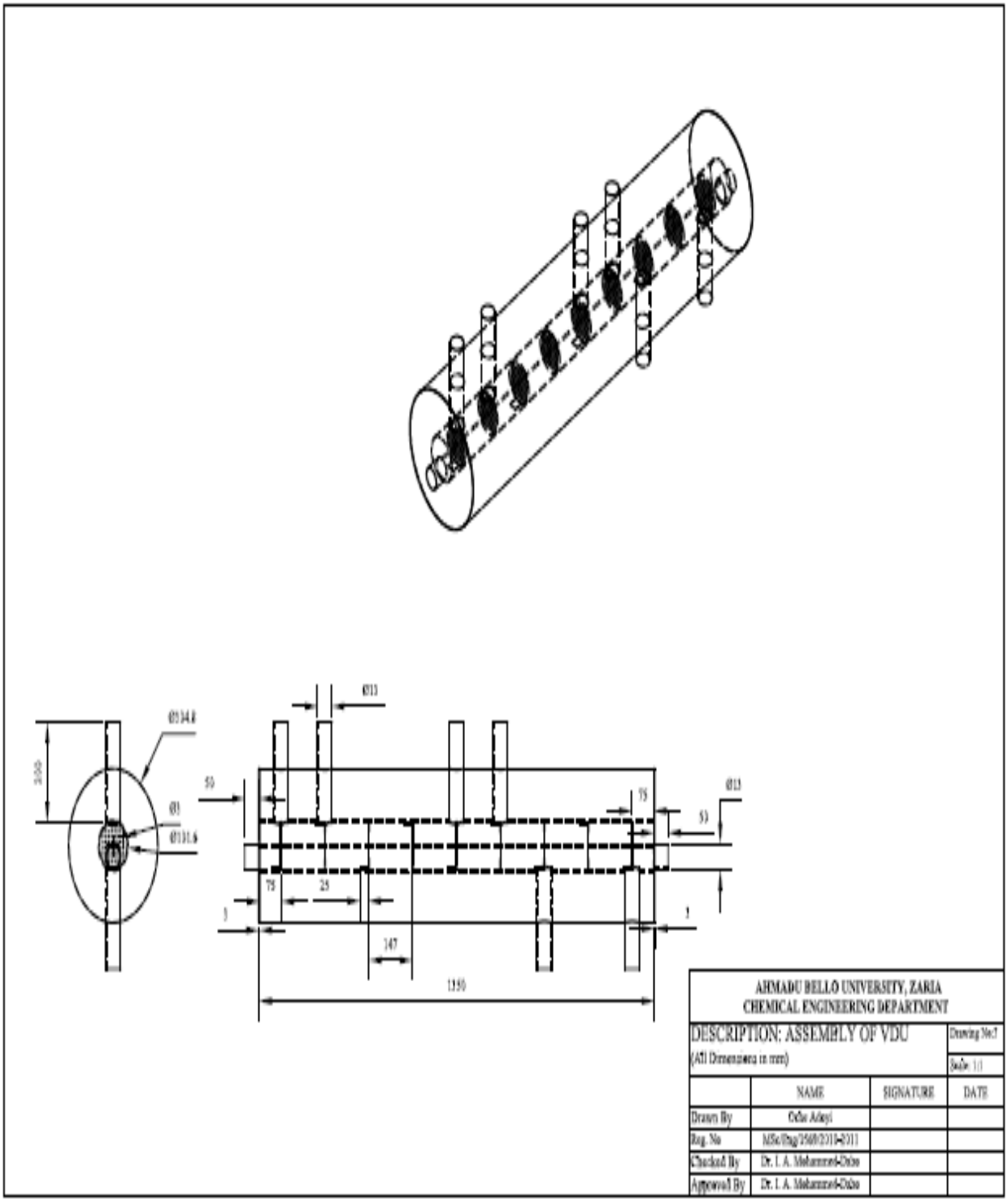


Figure 5.2: Vacuum Distillation Unit Working Drawing

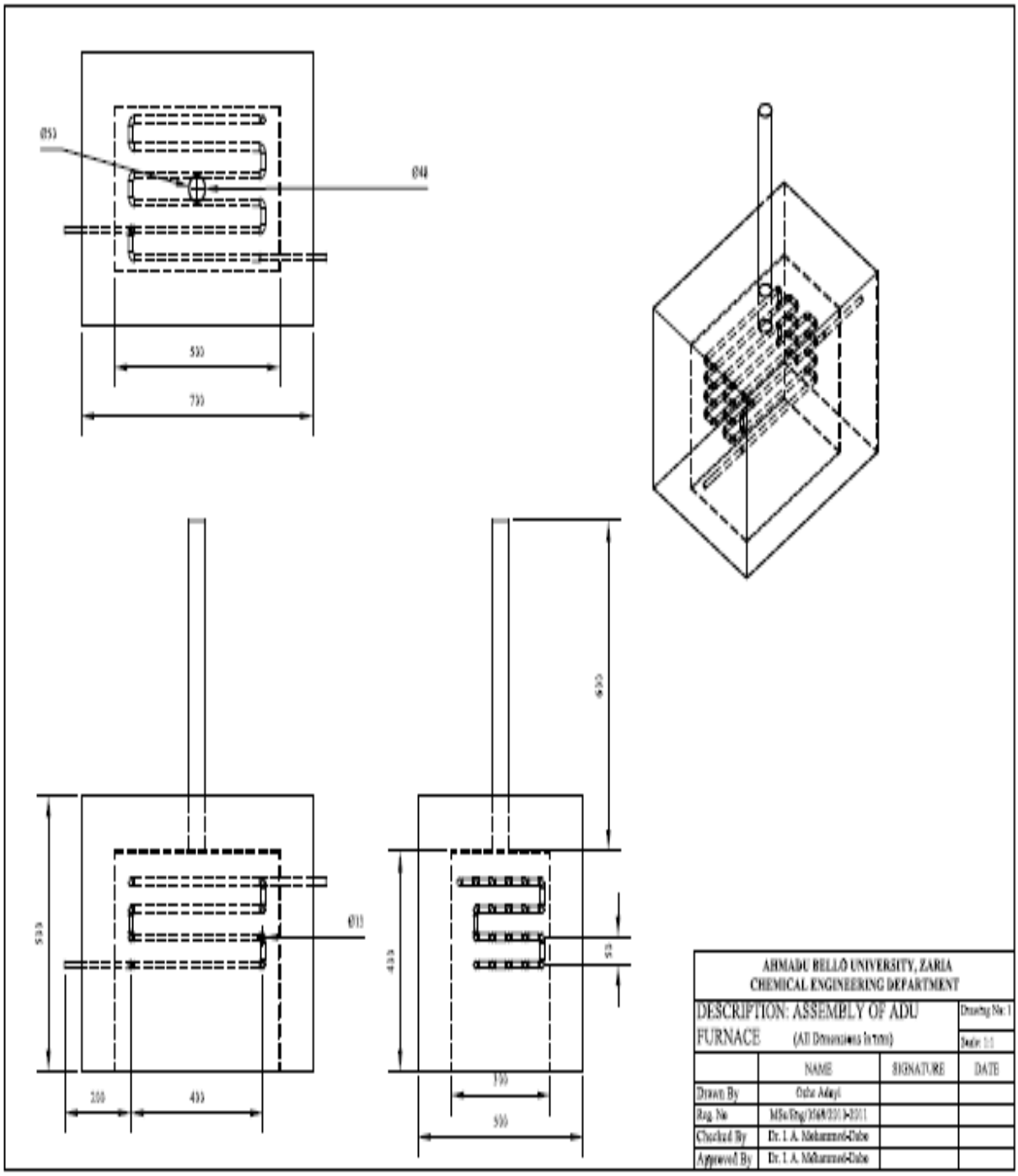


Figure 5.3 : ADU Furnace Working Drawing

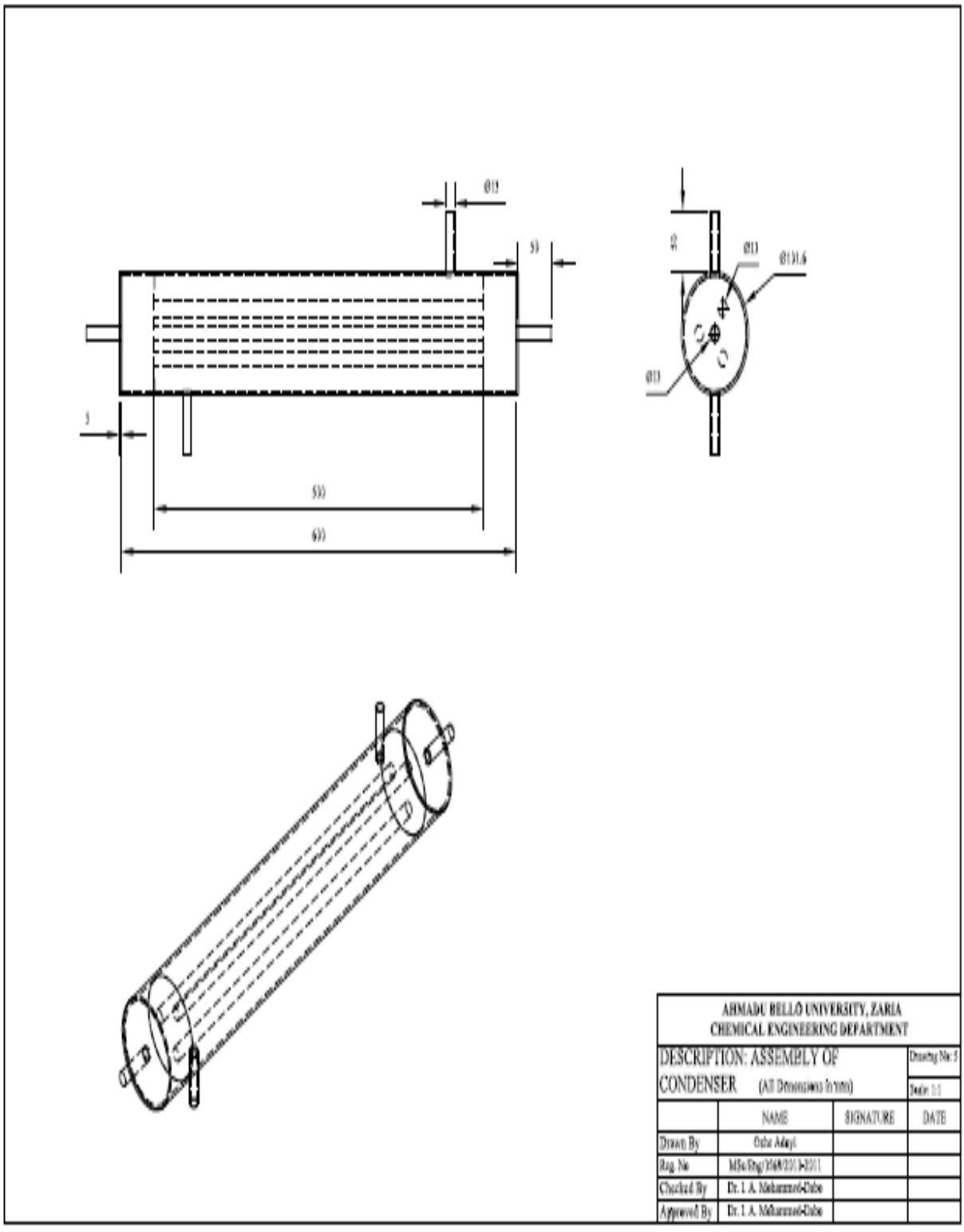


Figure 5.4: Condenser Working Drawing

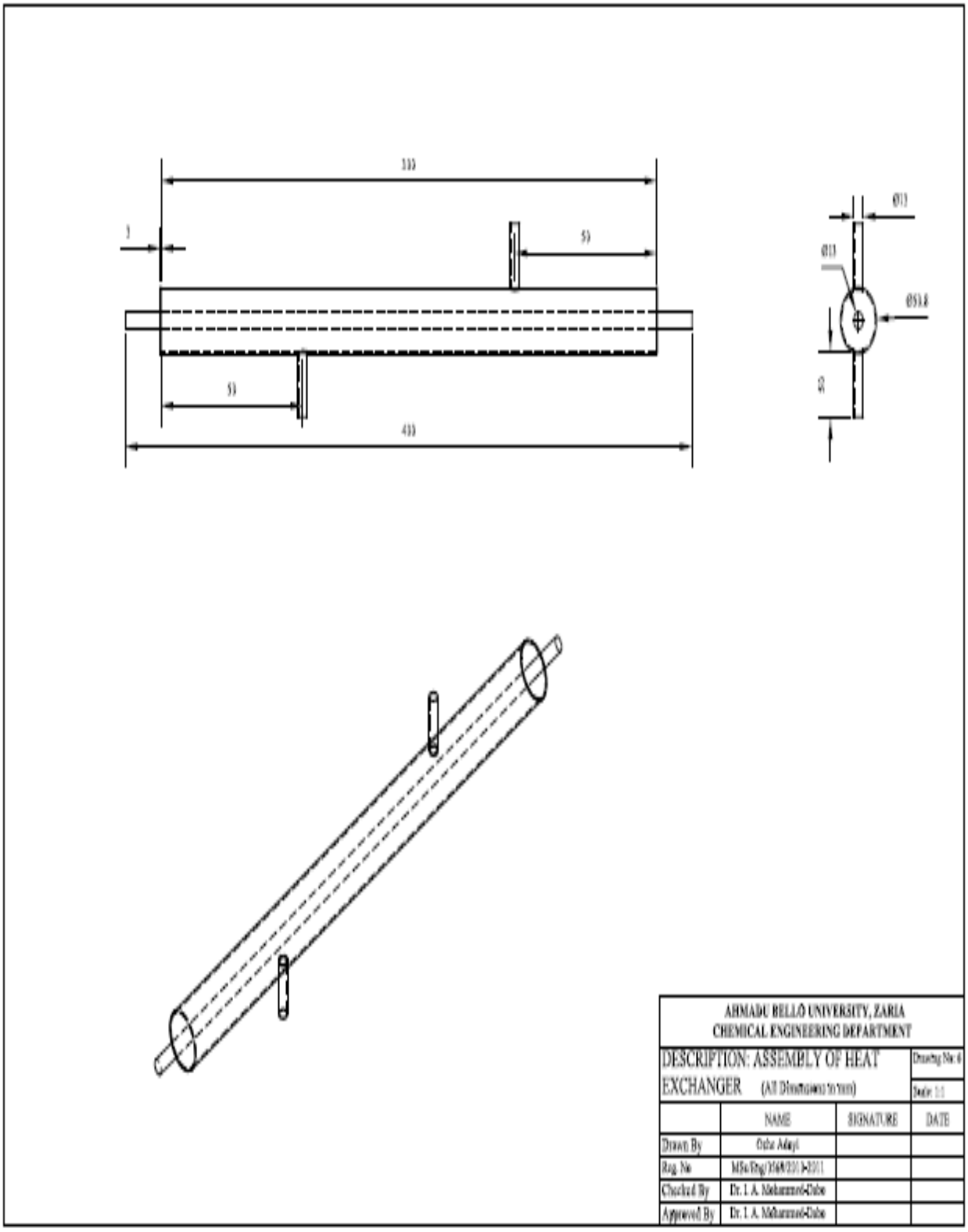


Figure 5.5: Heat Exchanger Working Drawing

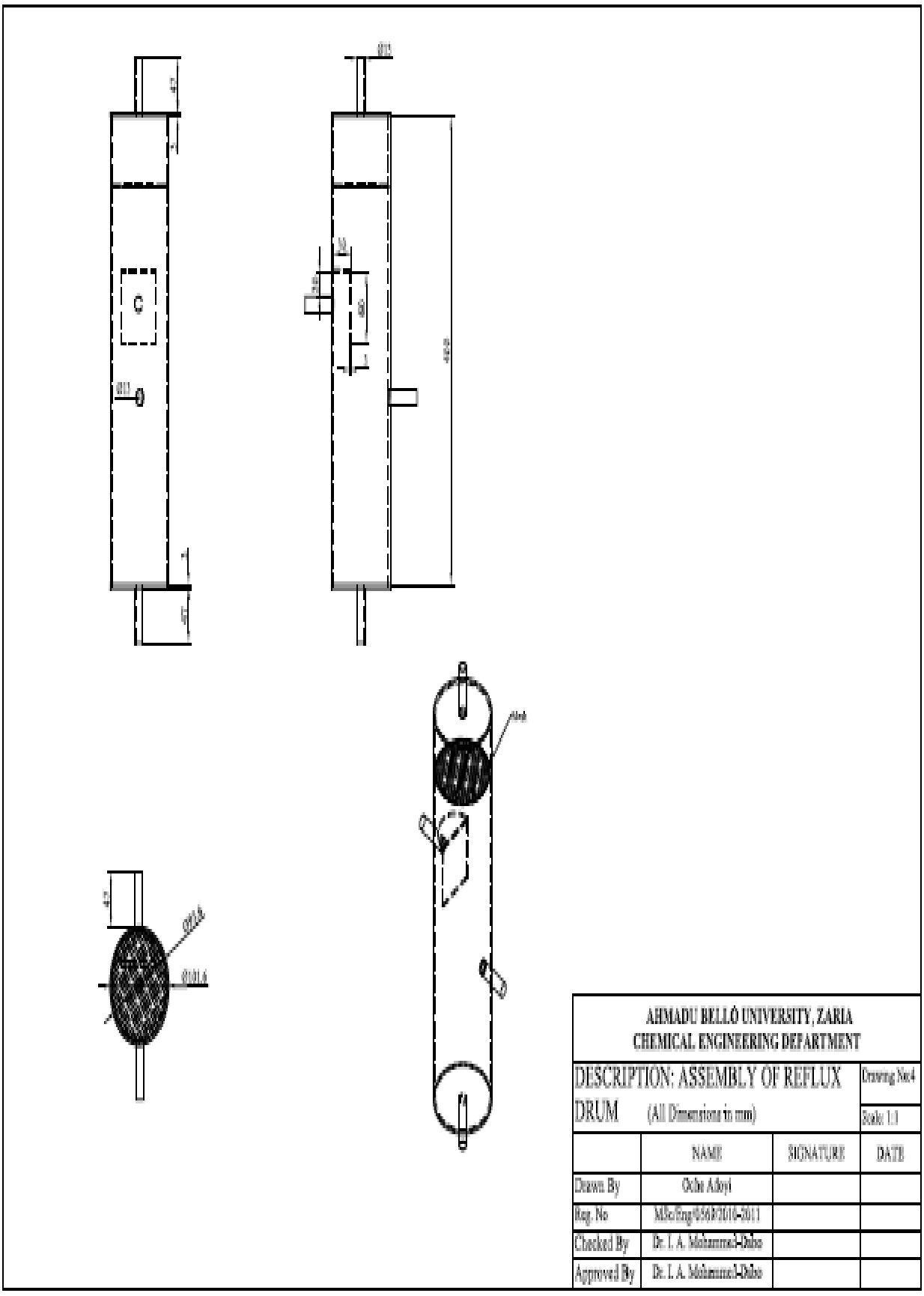


Figure 5.6: Reflux Drum Working Drawing

5.2 Layout Drawings for Mini Refinery

A proposed layout for the mini refinery is presented in Figure 5.7 and Figure 5.8. Figure 5.7 is a top view of the layout with dimensioning while Figure 5.8 is side view of the same layout. The flammable crude and products are appropriately spaced away from the heat areas (furnaces, columns and steam generator). Levels of the units were appropriately positioned to maximize flow by gravity. Drawing was done with Google Sketch-Up Software.

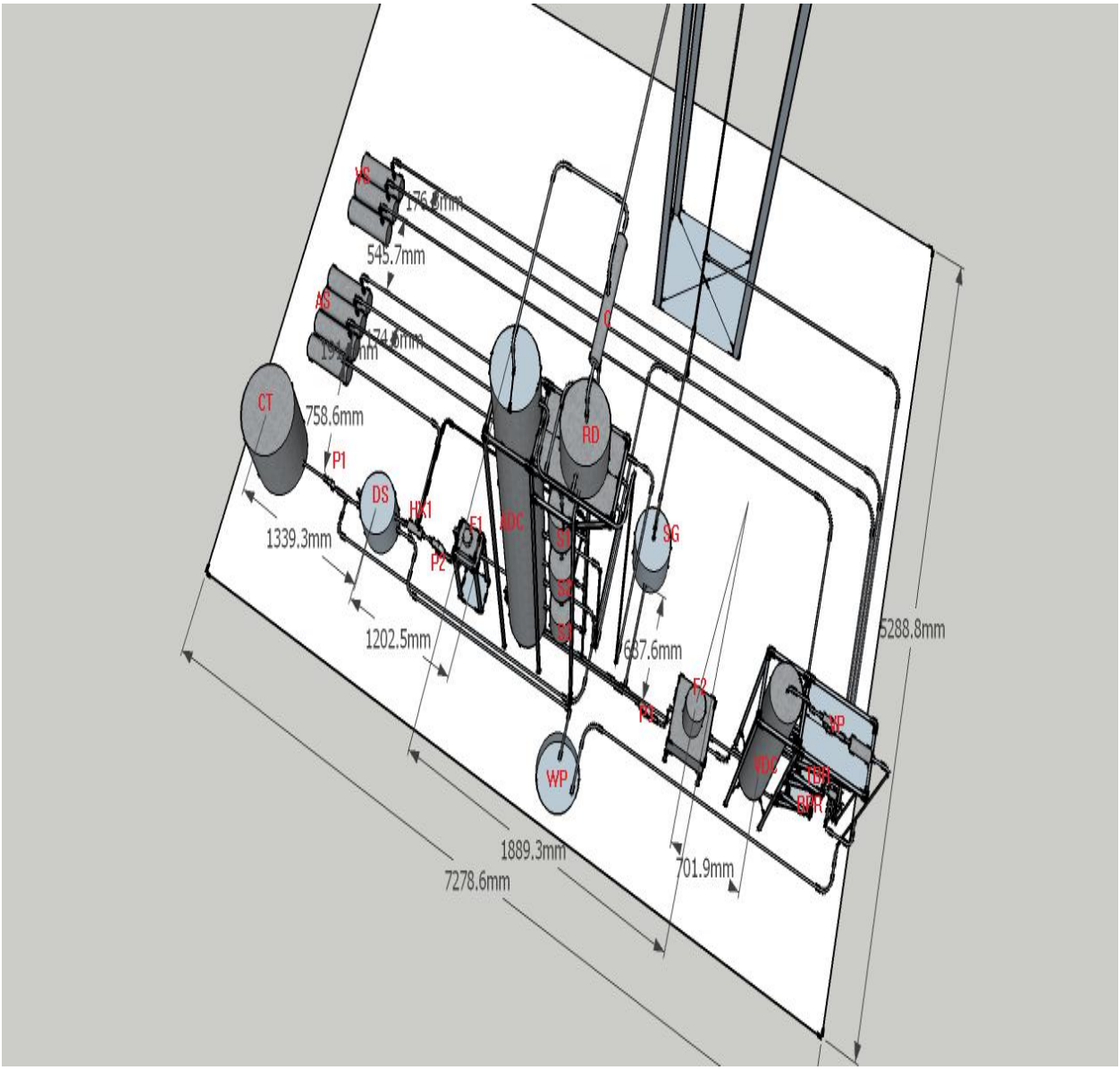


Figure 5.7: Mini-Refinery Layout

Key for Figure 5.7

BPR	Bottom Pump Around	SG	Steam Generator	DS	Desalter
VS	Vacuum Product Storage	VP	Vacuum Pump	P1,2,3	Pumps
ADC	Atmospheric Distillation Column	WP	Waste water Pit	S1,2,3	Strippers
TBR	Top Pump Around	CT	Crude oil storage	F1,2	Furnaces
VDC	Vacuum Distillation Column	RD	Reflux Drum	C	Condenser
AS	Atmospheric Product Storage	HX1	Heat exchanger		

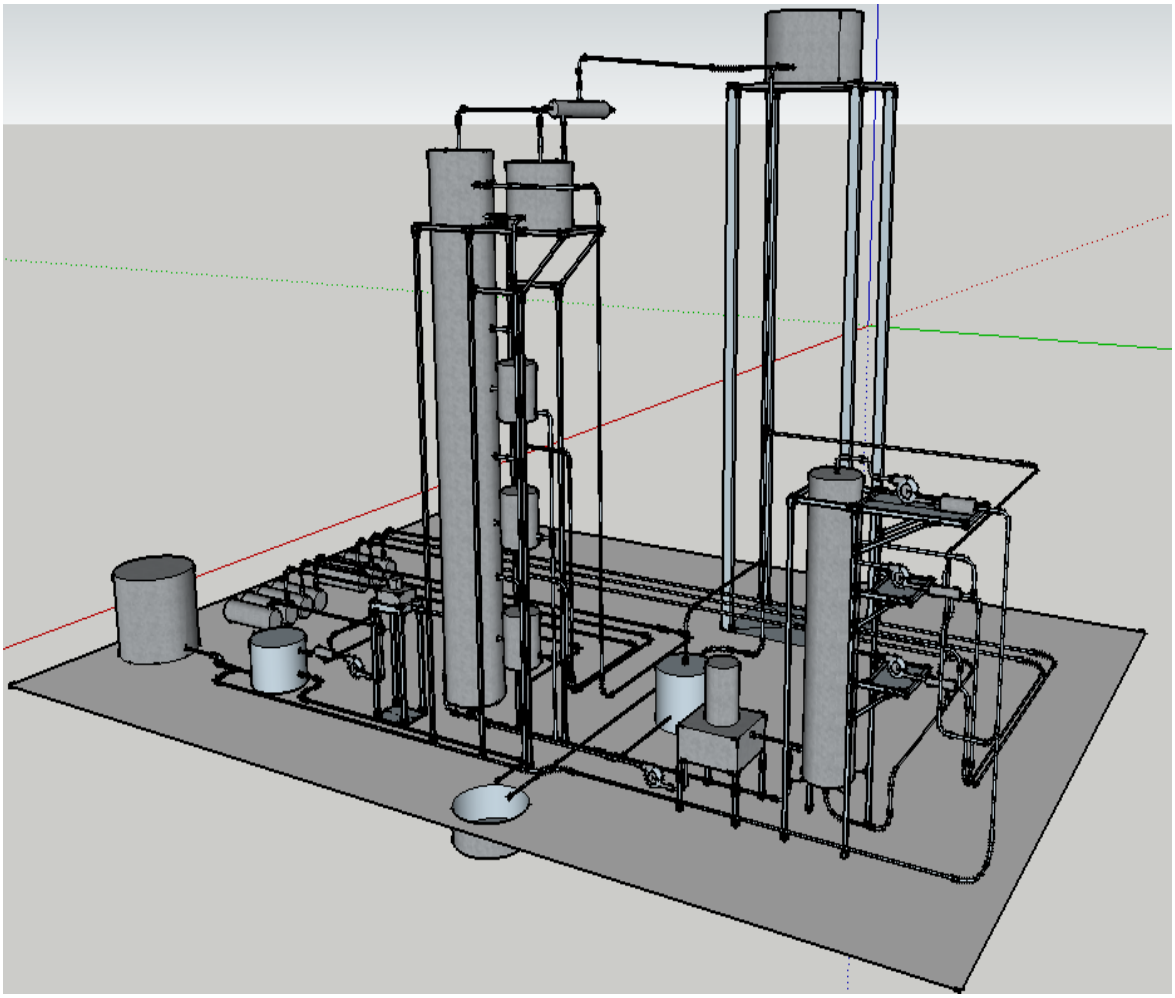


Figure 5.8: 3D drawing for the Mini-Refinery

5.3 Fabrication

Fabrication was done at Hanigha Company, Kakuri-Kaduna. The supervisory team from Hanigha Company on the work was made up of the managing director, a consultant and two company supervisors. Over fifteen welders, iron benders and machine operators were utilized for the work. Very close supervision was also done by project student, project supervisor, technical staffs from Department of Chemical Engineering and Mini Refinery Development Research team members. The working drawings and detail of materials required were taken to fabricator. Stainless steel was estimated to cost well above budgeted project capital. Mild steel was used in place of stainless steel. Cylindrical vessels in design were approximated to nearest standard pipe size in the market as it cheaper than fabricating all vessels. Lengths of pipes used in each component were summed up to estimate quantity of pipes required and areas of metal sheets to be used in each component was summed up to estimate quantity of metal sheets required. All materials obtained by fabricator met specification. Fabrication was done with strict adherence to technical and design standard. Fabrications that did not meet standards were discarded and redone better as shown in Plate 5.1 to Plate 5.26.

Air pressure test was carried out on all components to check their ability to withstand pressurized fluid when in operation. Here is a brief description of the test procedure:

I. The mini refinery component is pipe with a rubber tube to an air compressor.
II. All other openings on component are tightly blocked with plugs.
III. The air compressor is then turned on to supply pressurized air.
IV. Each weld point is checked for leakages and failure by applying soapy water at weld point.
V. Formation of bubble as soapy water is applied indicate leakage, the point is marked for proper welding after test.

5.3.1 Furnace Tubes

A mild steel pipe of thickness 4 mm, diameter 13 mm and length 40 cm was used for tubes. In first trial, tubes were joined by welding elbows. Too many weld joint were available with a number of them failing air pressure test.

Second trial for furnace was a better technique developed to take care of the short coming of first trial. The joints were made by bending the length of pipe to get tube train. But the second trial had collapse at some bends which failed the air pressure test.

The final trial passed air pressure test as it had no leakage at weld points and no collapse at bends. So it was accepted.

5.3.1.1 Furnace ADU Tubes Fabrication

The first, second and final fabrication of ADU Tubes are shown in Plate I – III.



Plate I: First trial – ADU Furnace



Plate II: Second trial – ADU Furnace tube



Plate III: Final trial – ADU Furnace tube

5.3.1.2 FURNACE VDU TUBES FABRICATION

The first, second and final fabrication of VDU Tubes are shown in Plate IV and V.



Plate IV: First trial – VDU furnace



Plate V: Final trial – VDU furnace

5.3.2 Furnace Chamber

First trial was modified to second/final trial based on the fact that the latter will perform better than the former with respect to circulation of hot air for heating purpose as shown in Plate VI and VII.



Plate VI: First trial – Furnace Chamber



Plate VII: Final trial – Furnace Chamber

Interior of the furnace box is shown in Plate VIII



Plate VIII: Couple Furnace interior view

The furnace boxes were tightly lagged all round with mineral wool of thickness 5 mm as shown in Plate IX.



Plate IX: Furnace chamber insulation

5.3.3 Distillation Columns

A. Column Sieves Trials

First column sieve trial were plates cut by hand cutter with hole drills of 3 mm as shown in Plate X. the second trial still had hole diameter 3 mm but it was cut by machine giving it a well rounded edge as shown in Plate XI.

The 3 mm hole diameter was discovered at a later stage to be outside what was specified in design. 3rd/final trial had hole diameter 2 mm as specified in design and was machine cut and so was accepted as shown in Plate XII.

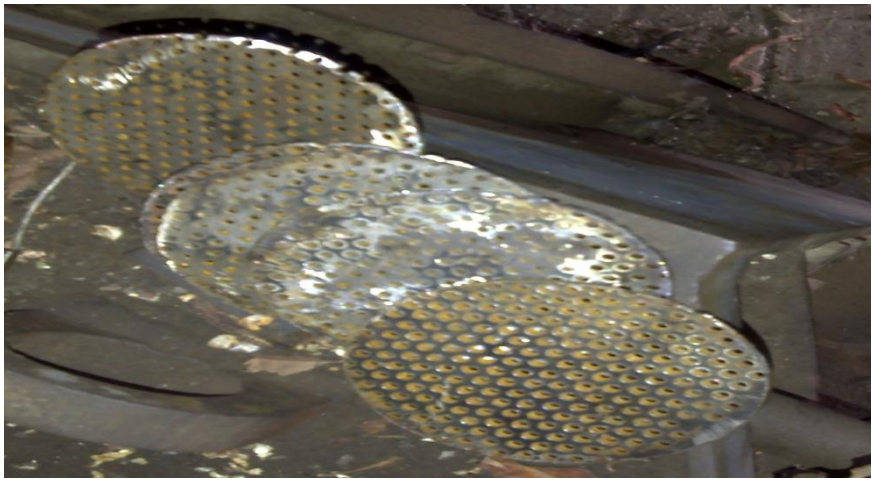


Plate X: 1st trial – Column sieves

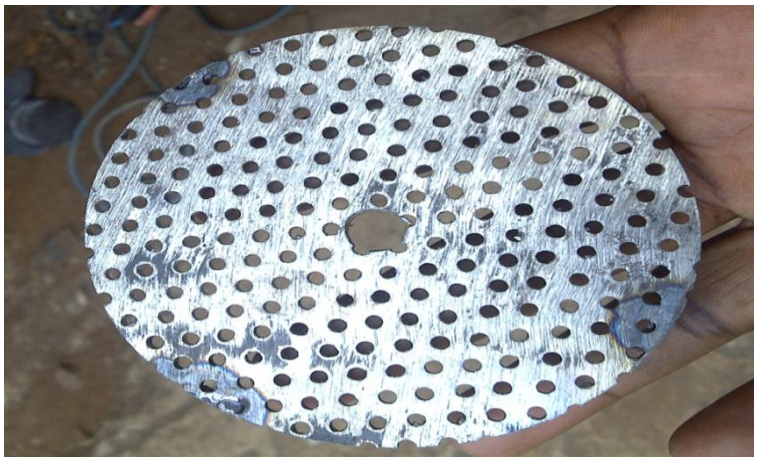


Plate XI: 2nd trial – Column sieves

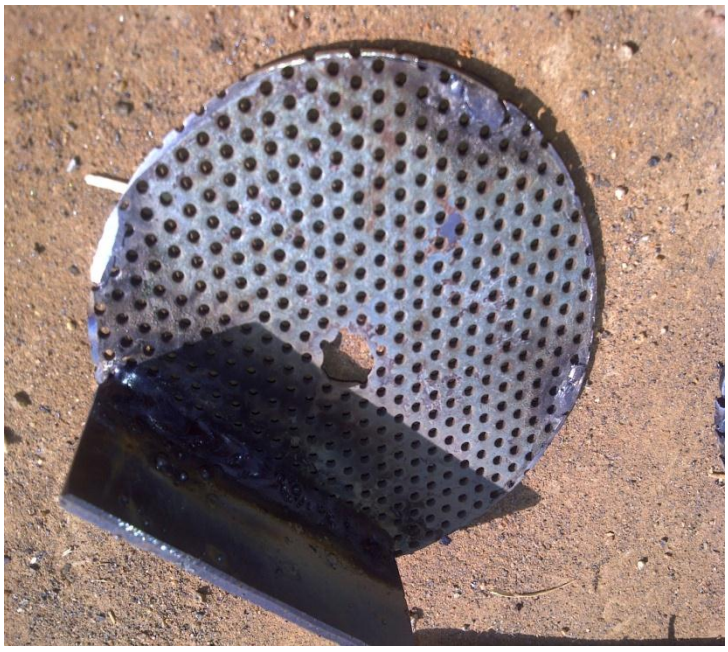


Plate XII: Final trial – Column sieves

The trays mounted on the spacer as to be inserted into distillation column is shown in Plate XIII.



Plate XIII: Mounted column sieves

The distillation column after being assembled is insulated with mineral wool as shown in Plate XIV.



Plate XIV: Distillation Column Insulation

Upper part of Distillation Column is shown in Plate XV.



Plate XV: Upper Part of Coupled Distillation Column

HEAT EXCHANGERS

Part of heat exchanger were machine cut as shown in Plate XVI and assembled as shown in Plate XVII.



Plate XVI: Heat Exchanger Components



Plate XVII: Coupled Heat Exchanger

Interior of condenser is shown in Plate XVIII.



Plate XVIII: Interior of Condenser

Finished Fabrication Works

Plate XIX to Plate XXV shows finished fabrication works for each of the components and labelling of process stream on them.



Plate XIX: Fabricated Atmospheric Distillation Column

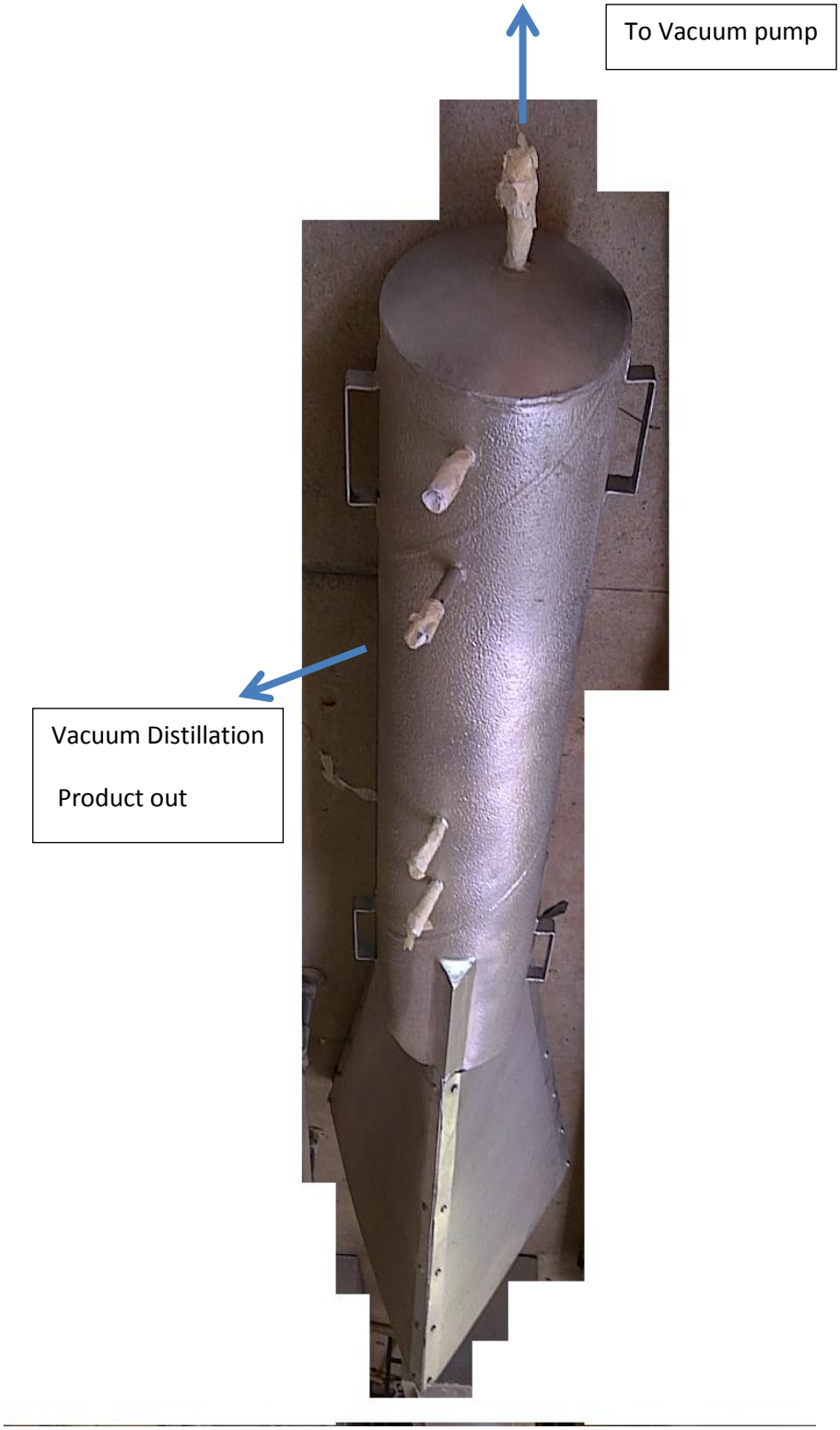


Plate XX: Fabricated Vacuum Distillation Column

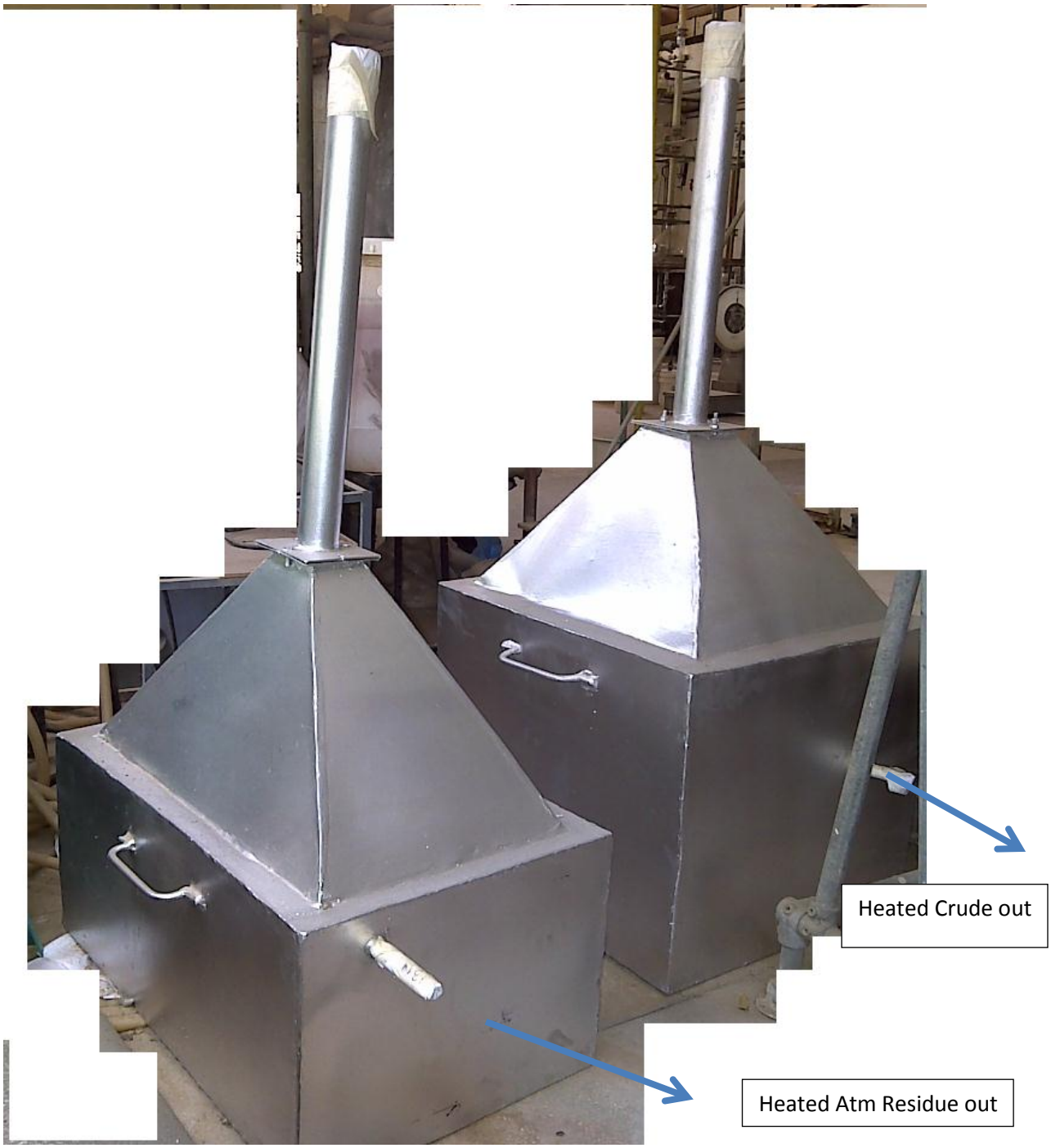


Plate XXI: Fabricated ADU and VDU Furnaces

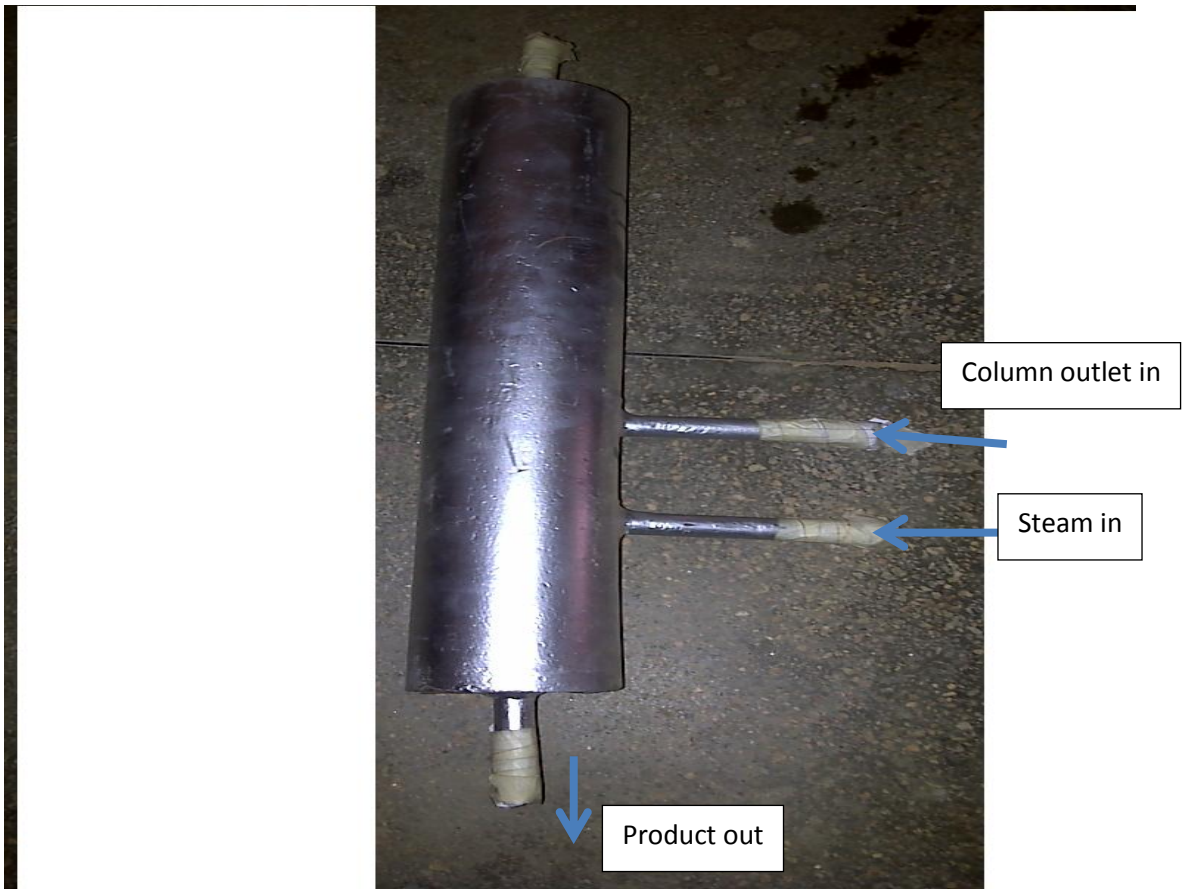


Plate XXII: Fabricated Stripper

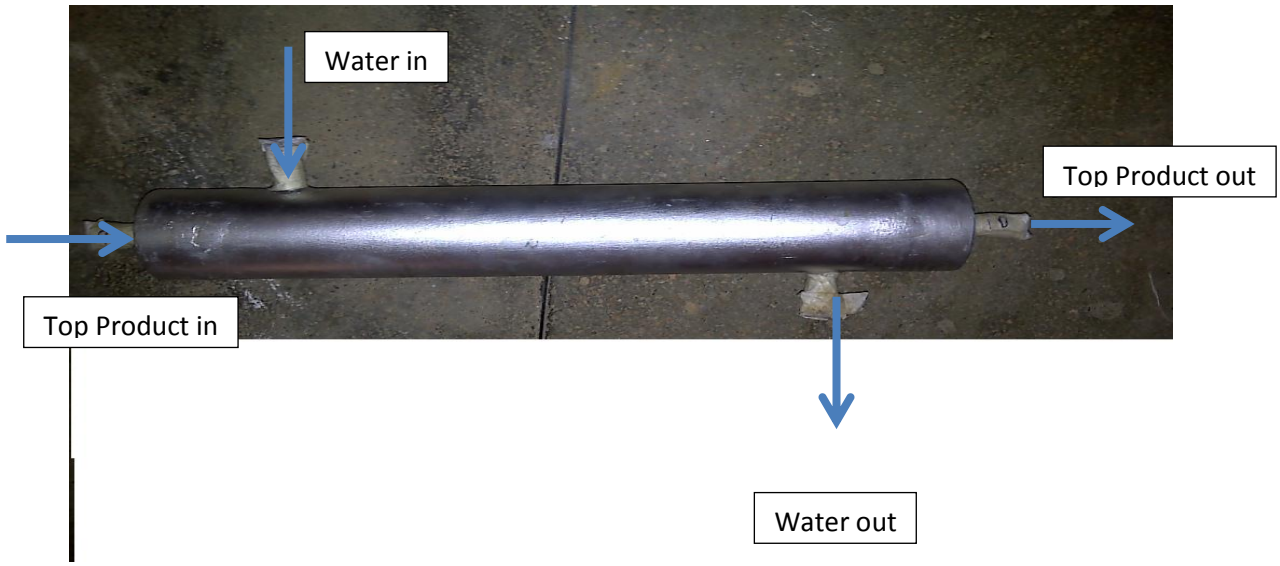


Plate XXIII: Fabricated condenser

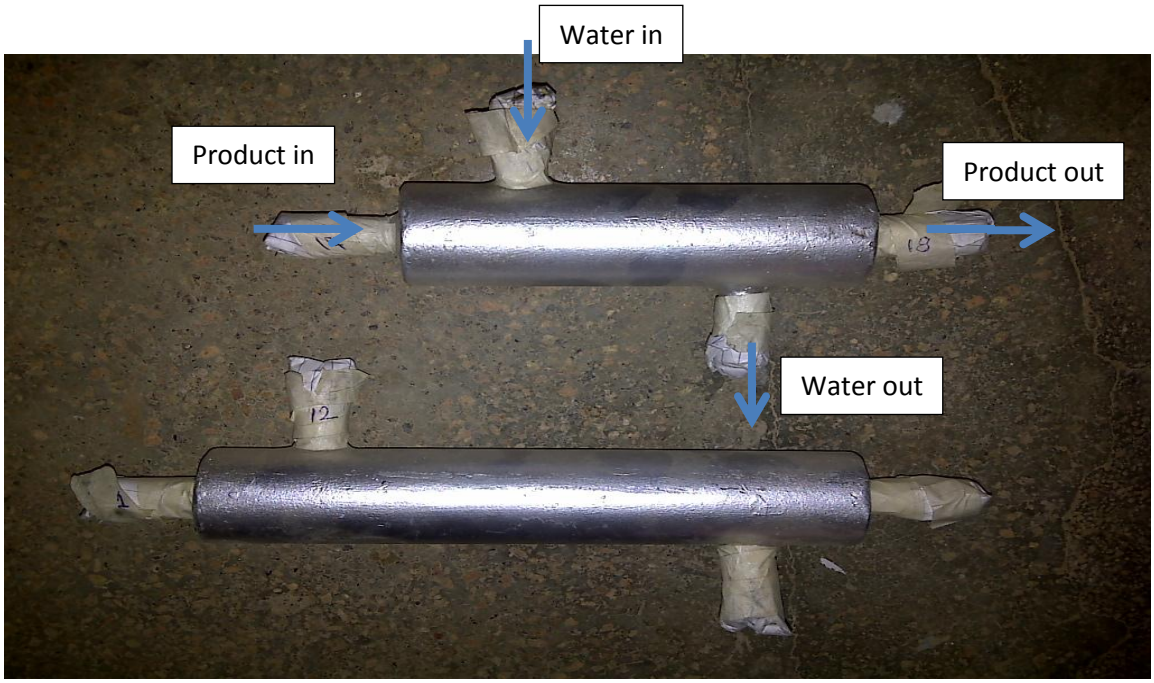


Plate XXIV: Fabricated Heat Exchangers

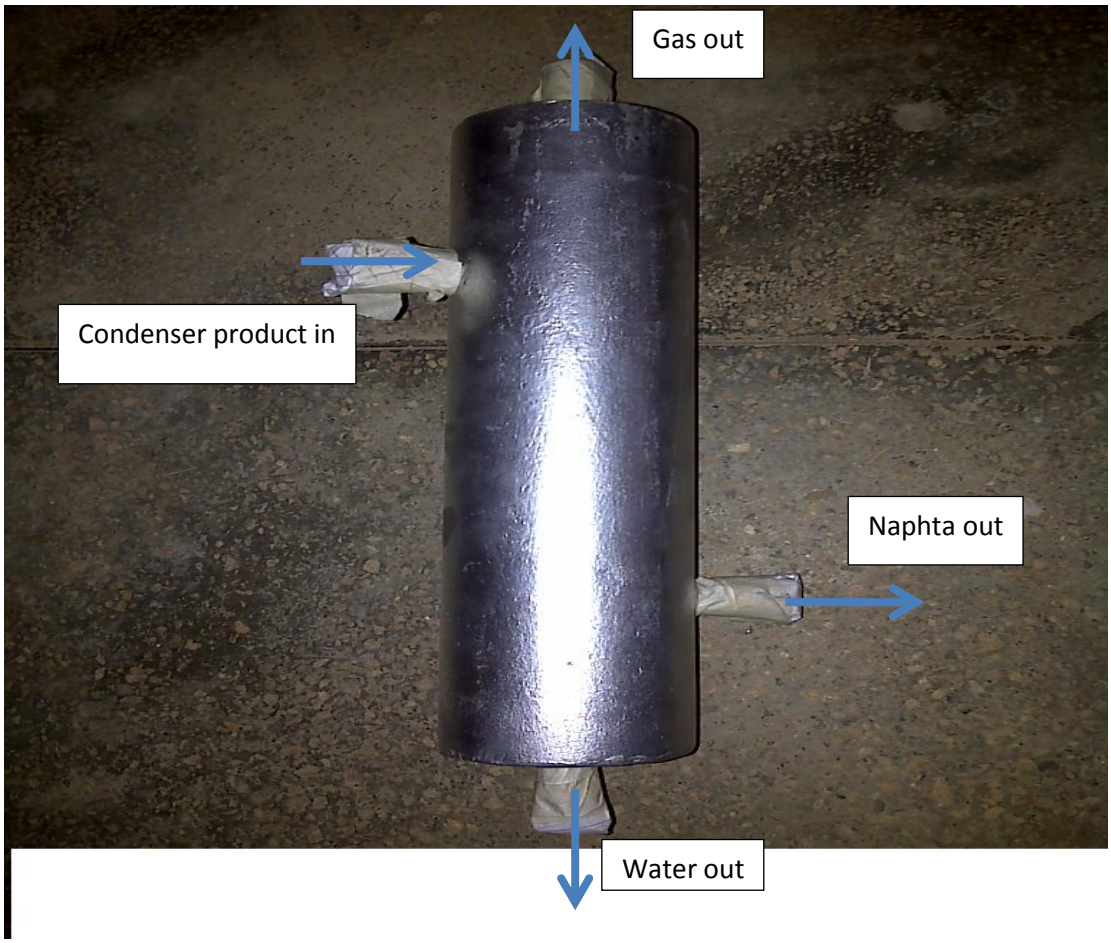


Plate XXV: Fabricated Reflux Drum

CHAPTER SIX

6.0 CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusions

The following are conclusions made:

A. For fractionating 1 bbl/day of Escravos crude:

I. Required number of trays for ADU, VDU and Stripper columns were found to be 20, 9 and 4 trays respectively.

II. For ADU, drawoff trays (stage) for heavy naphtha, kerosene and diesel are tray 6, 13 and 17 respectively and feed stage is tray 18.

III. For VDU, drawoff trays for LVGO and HVGO are tray 2 and 6 and feed stage is 7.

IV. Internal diameter of ADU, VDU and Stripper columns are 7.5cm, 9.9cm and 2cm respectively.

V. ADU furnace was determined to be a double section box furnace with 12 tubes in convection section and 5 tubes in radiation section while VDU furnace was determined to be 4 tubes in radiation section only. All tubes are 0.4m long with diameter of 5mm.

VI. ADU top product condenser required is a shell and tube heat exchanger with 3 tubes of length 0.5m and diameter 5mm while crude-AGO, Top pump around and bottom pump around heat exchangers are double pipe heat exchanger with tube lengths 0.3m, 0.2m and 0.2m respectively.

VII. Reflux drum required are vertical separators of height and diameter; 32cm and 5cm respectively.

B. The components for the mini refinery have successfully been fabricated.

6.2 Recommendations

The following are recommendations made:

I. The components for the mini refinery should be installed and piped with full consideration of mechanical and structural stability.

II. An effective control system should be designed, constructed and installed for Mini refinery.

III. Test run on Mini refinery should be carried out at design operating parameters specified in this work. Changes can only be carried out after proper simulation of the process with appropriate software.

REFERENCES

Akpa, J. G. & Umuze O. D. (2013), Simulation of a Multi-component Crude Distillation Column.American Journal of Scientific and Industrial Research, 4(4), 366-377.

Aspen Plus Guide (2005), Aspen Plus 2004.1: Getting Started Modelling Petroleum Processes. Aspen Plus, Inc. Cambridge, USA.

Bagajewicz, M. & Ji, S. (2001), Rigorious Procedure for the Design of Conventional Atmospheric Crude Fractionation Units. Part I: Targeting: Ind. Eng. Chem. Res. 40.

Burklin, C. E. (1977), Revision Of Emission Factors for Petroleum Refining, EPA-450/3-77-030, U. S. Environmental Protection Agency, Research Triangle Park, NC.

Eman, E. A. (2013), Clays as Catalysts in Petroleum Refining Industry. ARPN Journal of Science and Technology, 3(4).

Ji, S. & Bagajewicz, M. (2002), Design of Crude Distillation Plants with Vacuum Units. I. Targeting. Ind. Eng. Chem. Res., 41, 6094-6099.

Kamel, D., Mamdouh, G. & Fatma, A. (2013), New Retrofit Approach for Optimisation and Modification for a Crude Oil Distillation System.Chemical Engineering Transactions, 35, 1363-1368 DOI:10.3303/CET1335227

Lekan, T. P., Gutti, B. & Alfred A. S. (2013), Expert System Design and Control of Crude oil Distillation Column of a Nigerian Refinery using Artificial Neural Network Model. IJRRAS 15 (3), 337-346.

Lopez, D. C., Luis, J. H., Leonardo, A. & Jaime, F. V. (2009), Optimization Model of a System of Crude Oil Distillation Units with Heat Integration and Metamodeling. Ciencia, Tecnología y Futuro, 3(5), 159-174.

Manasseh, S. (2001), Design of a Medium size Refinery to Process 20,000 barrels per stream day. Chemical Engineering ABU Zaria.

Massimiliano, E. et al (2008), Energy Saving in a Crude Distillation Unit by a Preflash Implementation. Elsevier Journal, Applied Thermal Engineering 29 (2009) 1642–1647.

Maxwell, J. B. (1950), Data book on hydrocarbons, 9th Edition, Von Nostrand Reinhold Company, USA.

Mohammed, A. (2002), Design of a Medium Size Refinery II: Upgrading of a Hydro-skimming Plant. Chemical Engineering ABU Zaria.

Nkaginieme, U. (2005), The Challenges of Building a New Refinery in Nigeria with Limited Energy Infrastructure and Regulated Petroleum Products Market, Totalsupport Energy Group, Africa Session, Forum 22 paper.

Nuhu, M., Yusuf, A. M., Souleymane, M. A., Sukairaji, A., Bilal, S. & Onukak, I. E. (2013), EnergyAnalysis of CrudeDistillationUnit (CDU) of N'DJAMENA Refinery (CHAD). International Journal of Scientific & Engineering Research, 4(5), 264-268.

Perry, R. H. & Green, D. W. (1997), Perry's Chemical Engineers' Handbook, 7th ed.: Mc Graw Hill Companies inc. New York.

Rao, B. (1990), Modern Petroleum Refining Processes, 1st ed. New Delhi India.

Sinnott, R. N. (2006), Coulson and Richardson's Chemical Engineering Series, Vol. 6, 4th Ed.:Elsevier New Delhi.

Timmerhaus, M. (1991), Plant Design and Economics for Chemical Engineers, 4th ed. Mc Graw Hill Companies inc. New York.

Ukaegbu, C. (2007), Leadership Fatalism and Underdevelopment in Nigeria: Imaginative Policymaking for Human Development, Philosophia Africana, 10(2).

Uppaluri, R. (2010), Refinery process Design (Lecture Notes): department of Chemical India. Pp1- 183

Vineet, K., Anuj, S., Indranil, R. C., Saibal, G. & Deoki, N. S. (2001), A Crude Distillation Unit Model Suitable for Online Applications. Elsevier Journal, Fuel Processing Technology 73 2001 1–21.

Crude Refining Africa: Way Forward. Retrieved August 25, 2013, from www.vanguardngr.com/2014/03/crude-refining-africa-way-forward/

Escravos Crude Assay. Retrieved July 2, 2014, from http://www.kerneloil.com/website/Assay/Africa/escravos_summary.pdf.

Illegal Refineries Detroyed in Delta State. Retrieved June 23, 2013, from http://dailypost.com.ng/2012/05/04/600-illegal-refineries-destroyed-in-delta-state.

Mini Refinery.Retrieved May 15, 2013, from http://www.minirefinery.com.

Nigerian Escravos Crude Assay. Retrieved July 2, 2014, from http://www.lavicorp.com/products/Nigeria/escravos.pdf.

Petroleum Refining Processes.Retrieved July 2, 2014, from http://chemengineering.wikispaces.com/Petroleum+refining+processes.

Reawakening the Ghost of Fuel Subsidy Removal.Retrieved August 25, 2013, from http://tribune.come.ng/index.php/editorial/46129-reawakening-the-ghost-of-fuel-subsidy-removal.

APPENDICES

APPENDIX A

A.0 Escravos Crude Assay

Table A.1: Crude and Crude product properties

PROPERTIES	WHOLE CRUDE	LIGHT GASOLINE(55-175^{0}F)	LIGHT NAPHTHA(175-300^{0}F)	HEAVY NAPHTHA(300-400^{0}F)	KEROSENE(400-500^{0}F)	ATM. GAS OIL(500-650^{0}F)	ATM RESDUUM(650^{0}F+)	LT VAC GAS OIL(650-800^{0}F)	HEAVY VAC GAS OIL(800-1050^{0}F)	VAC RESIDUUM(1050+^{0}F)
Gravity, ^{o}API	33.7	80	55.8	47.2	35.4	30.8	19.6	23.9	19.6	11.1
Specific Gravity	0.86	0.67	0.76	0.79	0.85	0.87	0.94	0.91	0.94	0.99
Mercaptan Sulfur, ppm	-	0.05	0.19	0.36	0.94	-	-	-	-	-
Octane Number, Research, Clear	-	75.4	66.4	-	-	-	-	-	-	-
Naphthenes,vol %	-	-	42.87	56.38	56.7	-	-	41.56	-	-
Aromatics,vol%	-	-	16.74	16.65	22.2	-	-		-	-
Paraffins,vol%								15.6	-	-
Smoke Point,mm	-	-	--	25.6	18.7	-	-	-	-	-
Freezing point,^{0}F	-	-	-	-	-49	-	-	-	-	-

Cetane Index	-	-	-	-	-	48	-	-	-	-
Cutter,vol% in Fuel oil	-	-	-	-	-	-	-	-	-	31.8
Fuel Oil Yield, vol%	-	-	-	-	-	-	-	-	-	9.75
Sulfur, wt%	0.16	-	-	0.02	0.06	0.13	0.31	0.23	0.31	0.47
Nitrogen,ppm	1190	-	-	-	-	89.9	3050	704	2550	8800
Hydrogen,wt%	-	-	--	-	-	-	-	12.56	12.19	11.12
Aniline Point, ^{0}F	-	-	-	-	-	-	-	172.2	188.6	-
Aniline Point, ^{0}C	-	-	-	-	-	-	-	77.9	87	-
PROPERTIES	WHOLE CRUDE	LIGHT GASOLINE(55-175^{0}F)	LIGHT NAPHTHA(175-300^{0}F)	HEAVY NAPHTHA(300-400^{0}F)	KEROSENE(400-500^{0}F)	ATM. GAS OIL(500-650^{0}F)	ATM RESDUUM(650^{0}F+)	LT VAC GAS OIL(650-800^{0}F)	HEAVY VAC GAS OIL(800-1050^{0}F)	VAC RESIDUUM(1050+^{0}F)
Pour Point ^{0}F	26.6	-	-	-	-	10.8	92.5	69.9	118	113.9
Pour Point ^{0}C	-3	-	-	-	-	-11.8	33.6	21.1	47.8	45.5
Acid Number, mg KOH/g	0.52	-	-	-	0.08	0.25	-	0.72	1.41	-
Back-Blended Acid,mgKOH/g	0.48	-	-	-	-	-	-	-	-	-

Viscosity@40^{0}C(104^{0}F),cSt	4.19	-	-	-	-	-	-	-	-	-
Viscosity@50^{0}C(122^{0}F),cSt	3.32	-	-	-	1.63	3.6	114	16.7	170	132000
Asphaltenes, C7,%	0.03	-	-	-	-	-	0.07	-	-	0.35
Nickel,ppm	4.55	-	-	-	-	-	11.9	-	-	56.8
Vanadium,ppm	0.51	-	-	-	-	-	1.34	-	-	6.57
MCR,wt%		-	-	-	-	-	3.7	-	-	16.7
Characterization Factor,K	11.74	-	-	-	-	11.59	11.79	11.62	11.82	11.9

Escravos Crude Oil Assay, www.lavicorp.com/products/Nigeria/escravos.pdf, 2008

Table A.2: TBP yields, vol%

TBP YIELDS,	Temperature ^{0}F	^{0}C	VOL %
Butane and lighter	-	-	1.00
Light Gasoline	55-175	12.8 – 79.4	4.36
Light Naphtha	175-300	79.4 -149	15.05
Heavy Naphtha	300-400	149 - 204.4	11.02
Kerosene	400-500	204.4 - 260	13.52
Atm. Gas Oil	500-650	260 - 343.3	20.18
Lt Vacuum Gas Oil	650-800	343.3 - 427	14.87
Hvy Vacuum Gas Oil	800-1050	427 - 565.6	13.36
Vacuum Residuum	1050^{0}F+	565.6+	6.64

Escravos Crude Oil Assay, www.lavicorp.com/products/Nigeria/escravos.pdf, 2008

Table A.3: Escravos Crude Assay

		Whole Crude	Light Naphtha	Medium Naphtha	Heavy Naphtha	Kero	Atm Gas Oil	Light VGO	Heavy VGO	Vacuum Resid	Atm Resid
TBP Temp At Start, °C			10	80	150	200	260	340	450	570	340
TBP Temp At End, °C			80	150	200	260	340	450	570	End	End
TBP Temp At Start, °F			55	175	300	400	500	650	850	1050	650
TBP Temp At End, °F			175	300	400	500	650	850	1050	End	End
Yield at Start, vol%			0.8	5.8	21.1	31.6	43.7	63.9	82.8	93.5	63.9
Yield at End, vol%			5.8	21.1	31.6	43.7	63.9	82.8	93.5	100.0	100.0
Yield of Cut (wt% of Crude)			3.9	14.0	9.8	11.9	20.6	20.1	11.7	7.6	39.3
Yield of Cut (vol% of Crude)			5.0	15.3	10.5	12.0	20.2	18.9	10.7	6.5	36.1
TBP Distillation, vol%	°C Start		10	80	140	200	260	340	450	570	340
	°C 5%		24	81	152	208	263	347	458	572	352
	°C 10%		29	91	156	211	267	352	462	578	362
	°C 30%		41	101	166	223	281	373	478	606	405
	°C 50%		60	118	178	234	298	397	497	636	449
	°C 70%		67	136	188	246	317	419	518	679	506
	°C 90%		72	143	199	256	334	442	547	760	627
	°C 95%		73	146	202	258	339	448	556	781	700
	°C End		80	150	210	260	340	450	570	End	End
TBP Distillation, vol%	°F Start		50	175	290	400	500	650	850	1050	650
	°F 5%		76	177	305	406	506	657	857	1062	665
	°F 10%		84	195	312	411	513	666	864	1073	683
	°F 30%		105	214	330	433	538	704	893	1123	781
	°F 50%		140	245	353	454	568	746	928	1177	840
	°F 70%		153	276	370	474	602	786	965	1254	943
	°F 90%		161	290	390	492	634	827	1017	1400	1161
	°F 95%		164	294	395	496	642	839	1032	1437	1282
	°F End		175	300	410	500	650	850	1050	End	End

Source: **www.kerneloil.com/website/Assay/Africa/escravos_summary.pdf**

Table A.4: Escravos Crude Assay continues

	Whole Crude	Light Naphtha	Medium Naphtha	Heavy Naphtha	Kero	Atm Gas Oil	Light VGO	Heavy VGO	Vacuum Resid	Atm Resid
TBP Temp At Start, °C	Start	10	80	150	200	260	340	450	570	340
TBP Temp At End, °C	End	80	150	200	260	340	450	570	End	End
TBP Temp At Start, °F	Start	55	175	300	400	500	650	850	1050	650
TBP Temp At End, °F	End	175	300	400	500	650	850	1050	End	End
Yield at Start, vol%		0.8	5.8	21.1	31.6	43.7	63.9	82.8	93.5	63.9
Yield at End, vol%		5.8	21.1	31.6	43.7	63.9	82.8	93.5	100.0	100.0
Yield of Cut (wt% of Crude)		3.9	14.0	9.8	11.9	20.6	20.1	11.7	7.6	39.3
Yield of Cut (vol% of Crude)		5.0	15.3	10.5	12.0	20.2	18.9	10.7	6.5	36.1
Gravity, °API	34.4	80.2	50.6	45.7	36.0	31.6	24.7	20.8	10.9	20.9
Specific Gravity	0.8528	0.6684	0.7769	0.7987	0.8447	0.8673	0.9060	0.9293	0.9934	0.9286
Sulfur, wt%	0.15	0.00	0.01	0.02	0.06	0.12	0.23	0.31	0.45	0.30
Mercaptan Sulfur, ppm		3	5	6	5	5	3	2		
Nitrogen, ppm	1268	0	0	2	8	106	880	3328	8991	3166
Hydrogen, wt%	13.1	16.1	14.0	13.8	13.4	13.0	12.7	12.3	11.3	12.3
Viscosity @ 40 °C (104 °F), cSt	4.75			1.00	2.08	4.59	32.1	531	5.19E+06	213
Viscosity @ 50 °C (122 °F), cSt	3.54			0.898	1.73	3.70	21.1	249	6.43E+05	115
Viscosity @ 100 °C (212 °F), cSt	1.26			0.590	0.867	1.66	4.96	22.3	2140	15.3
Viscosity @ 135 °C (275 °F), cSt	0.785			0.477	0.624	1.14	2.63	8.41	299	6.55
Freeze Point, °C				-72.000	-43.000	-4.000	34.0	53.0		
Freeze Point, °F				-97	-46	24	94	127		
Pour Point, °C	7			-74	-48	-8	29	47	53	35
Pour Point, °F	45			-102	-55	17	84	116	127	95
Smoke Point, mm (ASTM)				27	19	15	13			
Aniline Point, °C				51	60	69	84	92	94	88
Aniline Point, °F				123	140	156	184	198	202	191
Total Acid Number, mg KOH/g	0.53			0.0	0.1	0.3	0.9	2.7	1.7	1.6
Cetane Index, ASTM D976				33	39	46				
Diesel Index				56	50	50	45	41	22	40
Characterization Factor (K Factor)	11.7	12.6	11.5	11.7	11.5	11.6	11.7	12.0	12.0	11.9
Research Octane Number, Clear		70.4	64.6	44.9						
Motor Octane Number, Clear		67.9	62.1							
Paraffins, vol%		80.1	31.2	41.1	21.9	29.4	24.2			
Naphthenes, vol%		19.9	54.8	38.2	55.5	44.0	42.7			
Aromatics, vol%		0.0	14.0	20.7	22.6	26.5	33.1			
Thiophenes, vol%										
Molecular Weight	209	101	112	143	172	218	313	454	728	375
Gross Heating Value, MM BTU/bbl	5.85	4.87	5.45	5.58	5.81	5.92	6.11	6.23	6.50	6.22
Gross Heating Value, kcal/kg	10900	11550	11120	11100	10930	10850	10700	10640	10400	10630
Gross Heating Value, MJ/kg	45.6	48.3	46.5	46.5	45.7	45.4	44.8	44.5	43.5	44.5
Heptane Asphaltenes, wt%	0.0								0.4	0.1
Micro Carbon Residue, wt%	1.4								18.0	3.5
Ramsbottom Carbon, wt%	1.1								14.7	2.8
Vanadium, ppm	1								7	1
Nickel, ppm	5								64	12
Iron, ppm	0								5	1

Source: **www.kerneloil.com/website/Assay/Africa/escravos_summary.pdf**

Table A.5: Crude TBP Temperature data

Vol.%	$T_{TBP}(^0F)$
0	-30
10	217
30	414
50	593
70	800
90	1056
100	1357

Derived from: Perry

APPENDIX B

B.0 Scale up Analysis

Basis: 10 bbl/day.

Configuration: no pump arounds, no side stippers and no steam introduced into column.

TARGET (from Crude Assay)

Stream	Vol %	Flow rate (bbl/day)	Specific gravity	Average Molecular Weight
LNAPH	20.41	2.04	0.74	109
HNAPH	11.02	1.10	0.79	143
KEROSENE	13.52	1.35	0.85	172
AGO	20.18	2.02	0.87	218
RESIDUE	34.87	3.49	0.94	375
CRUDE	100	10	0.86	209

ASPEN PLUS RESULT

Stream	Flow rate (bbl/day)	Flow rate (lit/hr)	Specific gravity	Average Molecular Weight
LNAPH	2.1	13.91	0.74	92
HNAPH	1.1	7.29	0.80	158
KEROSENE	1.4	9.28	0.83	204
AGO	1.5	9.94	0.88	288
RESIDUE	4.0	26.50	0.94	416
CRUDE	10	66.25	0.86	209

OTHER RESULTS from Aspen Plus

Condenser duty = 20012.20 Btu/hr

Distillate rate = 13.6 lit/hr

Reflux rate = 41.95 lit/hr

Reflux ratio = 3.09

Condenser pressure = 15 psi or 103421.36 Pa

Temperature at condenser = 38^0C

Furnace duty =49259.57 Btu/hr or 12413.38 kcal/hr

Furnace pressure = 24.18 psi or 166715.23 Pa

Temperature at furnace = 350^0C

ASPEN PLUS SIZING RESULT

Tray spacing = 6 in

Column diameter = 0.733 ft or 22.34 cm

Downcomer area/column area = 0.1

Side weir length = 0.5325 ft or 16.23 cm

Hole diameter = 0.042 ft

Sieve hole area to active area = 0.12

Wier height = 1.5 in

Furnace calculation

Heat duty = 5043 kcal/hr

Total heat with 18% losses = 5043 + (0.18 x 5043)

= 5 950.74 kcal/hr

Heat value of fuel = 13 000kcal/kg

Amount of fuel required = 5 950.74/13000

= 0.458 kg/hr

The furnace consists of convection and radiation section. Calculations are carried out for sections separately.

Convective Section

Let 30% heat be absorbed in convection section.

Outlet temperature of crude in convection section = (350 – 190)(0.3)

$= 48^0C$

i.e. $48 + 190 = 238^0C$

Amount of fuel = 0.458 kg/hr

For excess air available, air/gas ratio = 25

Amount of air used = 0.458 x 25 = 11.45 kg/hr

Total gas = air used + fuel =11.45 + 0.458 kg/hr

= 11.9 kg/hr

For gas

$Qc = mC_p\Delta t$

1cal = 4.184J

Heat duty, Q = 5 950.74 kcal/hr x 4.184 = 24 898 kJ/hr

$0.3 \times 24\,898 = 11.9 \times 1.374 \times \Delta t$

$\Delta t = 457^0C$

Gas inlet temperature $t = 457 + 30 = 487^0C$

From countercurrent flow temperature curve

$\Delta t1 = 487 – 238 = 249^0C$

$\Delta t2 = 200^0C$ (Rao 1990)

LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$

= 224^0C

Ac = $\frac{Qc}{U(LMTD)} = \frac{0.3 \times 24\,898}{85.86 \times 273}$

= 0.389 m^2

PIPE DIAMETER SPECIFICATION

For carbon steel

d, optimum = 293 $G^{0.53}$ $\ell^{-0.37}$

G = 56.87 kg/hr

ℓ = 860 kg/m^3

d, opt = 293 x $\left(\frac{56.87}{3600}\right)^{0.53} 860^{-0.37}$

= 2.669 mm ≈3 mm

This is not a standard size pipe, 10 mm can be used.

Let L = 0.5m

D = 0.01m

No of tubes = $\frac{A}{\pi DL} = \frac{0.389}{\pi \times 0.01 \times 0.5}$ = 25 tubes

Radiation Section

For crude

$mC_p\Delta t = Q_R$

56.865 x 2.268 x (350 – 238) = R x 24 898

R = 0.58

$$R = \frac{1}{1+G\frac{\sqrt{\frac{Q}{\propto A}}}{S}}$$

Where constants s = 14200, $\propto$ = 0.88

$$0.58 = \frac{1}{1+25\frac{\sqrt{\frac{24\,898}{\propto A}}}{14200}}$$

$\propto$A = 0.147

A = N x D x space x L = N x 0.01 x 1.5 x 0.5 (space = 1.5D)

0.147 = 0.88 x N x 0.0075

N = 23 tubes

PUMP DESIGN

Velocity of crude = $\frac{10 \times 159 \times 4}{3600 \times 24 \times \pi \times 1000 \times (0.01)^2}$ = 0.23 m/s

Reynolds No. = $\frac{0.23 \times 0.01 \times 860}{0.01 \times 10^{-4}}$ = 2 x 10^6

Reynolds no. shows that flow is in turbulent region.

$P_C + P_R = \frac{4fL\ell V^2}{D}$

From chart, f = Fanning friction factor = 0.004

$P_C + P_R = \Delta P_f = \frac{4 \times 0.004 \times 0.5 \times (50+23) \times (0.23)^2}{0.01}$ = 3.1 N/m^2

Taking total energy balance for crude storage, pump and furnace

$g\Delta z + \Delta P/_{\varrho} - \Delta P_f/_{\varrho} - W = 0$

where Δz is difference in elevation,m

ΔP is difference in system pressure,N/m^2

W is work done by pump.

$g\Delta z$ and $\Delta P_f/_{\varrho}$ are negligible.

$-W = - \Delta P/_{\varrho}$

$-W = - (103421.36 - 166715.232)/860$

$= 77.2$ J/kg,

Power = (W x m)/η

M = mass flow rate = 56.87 kg/hr

Let η = efficiency = 0.7

Power for pump = (77.2 x 56.87)/0.7

= 6271.95 W = 6.3 kW

= 8.45 hp.

10 hp pump can be used.

Condenser Design

Condenser duty = Q = 3426.82 Btu/hr = 1004.3 W = 1 kW

For cooling water flow rate

$Q = mc\Delta T$

C = specific heat capacity of water = 4.18 kJ/kg^{0}C

Let cooling water be available at 30^{0}C and the temperature rise is to be limited to 10^{0}C.

Cooling water flow = $\frac{1}{(40-30)4.18}$ = 0.024 kg/s

From table 12.1 (Sinnott 2005),

Assume overall coefficient = 900 W/m^{20}C

We design for a horizontal shell and tube heat exchanger.

A = $\frac{Q}{U(LMTD)}$

LMTD = $\frac{\Delta t1 - \Delta t2}{2.303 log \frac{\Delta t1}{\Delta t2}}$

Δt1 = 117 – 40 = 77

Δt2 = 53 – 30 = 23

LMTD = $\frac{77 - 23}{2.303 log \frac{77}{23}}$

= 44.7^{0}C

A =$\frac{1004.3}{900(44.7)}$ = 0.025 m^2

Surface area of one tube = **π**DL

Take D = 5mm o.d.

L = 0.5 m

Surface area of one tube = **π** x 0.005 x 0.5 = 0.00785 m^2

Number of tubes = $\frac{0.25}{0.00785}$ = 32.

Calculation of actual column diameter

Let spacer diameter be 5 mm.

Cross sectional area of spacer = $\pi d^2/4$

$= \pi (0.005)^2/4$

$= 0.0000196\ m^2$

For main column,

Required area = $(0.2234)^2/4$

$= 0.0392\ m^2$

Actual Area = required area + 2 x area of spacer

= 0.039197 + 2 x 0.0000196

$= 0.03924\ m^2$

Actual diameter = $\sqrt{(4 \times \text{Actual area}/\pi)}$

$= \sqrt{(4 \times 0.03924/\pi)}$

= 0.2235 m

Number of holes by tray

Sieve hole area = 0.12 x Active area (Aspen Plus simulation)

Downcomer area/column area = 0.1

Column area = $0.039197\ m^2$

Downcomer area = 0.00392 m^2

Active area = Column area – 2 x downcomer area

= 0.0392 – 2 x 0.00392

= 0.03136 m^2

Sieve hole area = 0.12 x 0.03136

= 0.00376 m^2

Let plate hole size = 3 mm diameter

One hole area = $\boldsymbol{\pi}$ x $(0.003^2)/4$ = 0.00000707 m^2

Number of holes = $\frac{Sieve\ hole\ area}{one\ hole\ area}$ = $\frac{0.00376}{0.00000707}$ = 532 holes

APPENDIX C

C.0 Tables and Charts used for Manual Calculation

NOTE: ALL TABLES AND CHARTS WERE SOURCED FROM Uppaluri 2010

Table C.1: Hydrocarbon liquid enthalpy data for MEABP = 200°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0.49	0.41	0.33	0.24	0.16	0.07
20	8.12	8.34	8.56	8.78	9.00	9.22
40	17.98	18.29	18.60	18.91	19.22	19.54
60	27.90	28.32	28.74	29.16	29.58	30.00
80	37.59	38.15	38.71	39.27	39.82	40.38
100	47.38	48.16	48.94	49.71	50.49	51.27
120	57.74	58.76	59.79	60.81	61.83	62.85
140	68.52	69.61	70.70	71.79	72.88	73.96
160	78.81	80.25	81.69	83.13	84.56	86.00
180	88.86	90.77	92.68	94.60	96.51	98.42
200	99.60	101.54	103.47	105.40	107.33	109.27
220	110.37	112.46	114.55	116.64	118.73	120.82
240	122.05	124.37	126.70	129.03	131.36	133.68
260	133.69	136.21	138.72	141.23	143.75	146.26
280	144.74	147.54	150.35	153.15	155.95	158.75
300	156.43	159.66	162.90	166.13	169.36	172.59
320	168.98	172.47	175.95	179.43	182.91	186.39
340	180.43	184.26	188.09	191.92	195.75	199.59
360	192.85	197.02	201.18	205.34	209.51	213.67
380	207.06	211.34	215.62	219.91	224.19	228.47
400	220.58	225.00	229.42	233.84	238.26	242.67
420	234.17	238.93	243.69	248.45	253.21	257.97
440	247.78	253.05	258.32	263.59	268.86	274.13
460	262.39	268.04	273.68	279.33	284.98	290.62
480	277.45	283.60	289.75	295.91	302.06	308.21
500	293.22	300.01	306.80	313.59	320.38	327.17

Source: Maxwell, 1950

Table C.2: Hydrocarbon vapor enthalpy data for MEABP = 200°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	182.29	179.55	176.81	174.07	171.32	168.58
20	187.88	185.37	182.86	180.36	177.85	175.34
40	194.67	192.23	189.79	187.36	184.92	182.48
60	201.76	199.36	196.97	194.57	192.17	189.78
80	208.69	206.47	204.26	202.04	199.82	197.60
100	215.52	213.65	211.79	209.92	208.05	206.18
120	222.35	220.76	219.16	217.56	215.97	214.37
140	229.24	227.77	226.29	224.82	223.35	221.88
160	236.45	235.14	233.83	232.52	231.21	229.90
180	243.96	242.89	241.81	240.74	239.67	238.59
200	251.70	250.84	249.97	249.11	248.25	247.38
220	260.16	259.65	259.13	258.62	258.11	257.60
240	267.77	267.64	267.51	267.38	267.25	267.12
260	275.72	275.81	275.89	275.98	276.06	276.15
280	284.69	284.87	285.06	285.24	285.42	285.61
300	293.23	293.64	294.06	294.47	294.89	295.30
320	302.34	302.96	303.57	304.18	304.80	305.41
340	311.03	312.03	313.03	314.03	315.02	316.02
360	320.06	321.42	322.77	324.13	325.48	326.84
380	329.97	331.47	332.97	334.46	335.96	337.46
400	339.71	341.41	343.10	344.80	346.49	348.19
420	349.78	351.83	353.88	355.92	357.97	360.02
440	360.60	362.80	365.00	367.20	369.39	371.59
460	370.59	373.18	375.77	378.36	380.94	383.53
480	381.21	384.16	387.12	390.07	393.03	395.98
500	392.13	395.17	398.21	401.25	404.29	407.34
520	404.19	407.12	410.06	412.99	415.92	418.86
540	414.90	418.19	421.48	424.77	428.06	431.35
560	424.77	428.72	432.67	436.63	440.58	444.53
580	437.34	441.54	445.74	449.93	454.13	458.33
600	447.85	452.59	457.34	462.09	466.83	471.58
620	459.48	464.45	469.42	474.39	479.35	484.32
640	470.59	476.04	481.50	486.95	492.40	497.86
660	481.81	488.00	494.19	500.38	506.57	512.76
680	494.36	500.86	507.36	513.85	520.35	526.84
700	507.41	513.71	520.02	526.33	532.64	538.95
720	519.67	526.24	532.81	539.38	545.95	552.52
740	532.34	539.21	546.07	552.93	559.80	566.66
760	545.20	552.41	559.62	566.84	574.05	581.26
780	557.96	565.55	573.14	580.74	588.33	595.92
800	571.08	578.89	586.70	594.52	602.33	610.15
820	584.25	592.40	600.55	608.70	616.85	625.00
840	597.11	605.50	613.88	622.27	630.65	639.03
860	609.73	618.42	627.10	635.78	644.46	653.14
880	622.88	631.97	641.05	650.13	659.21	668.29
900	637.09	646.43	655.76	665.09	674.43	683.76
920	651.13	660.64	670.14	679.65	689.16	698.67
940	664.71	674.57	684.44	694.30	704.17	714.04
960	678.71	689.03	699.34	709.66	719.97	730.29
980	692.89	703.63	714.38	725.12	735.87	746.61
1000	706.48	717.70	728.92	740.14	751.35	762.57
1020	720.64	732.32	743.99	755.67	767.34	779.02
1040	734.09	746.28	758.46	770.64	782.83	795.01
1060	748.40	760.93	773.46	785.99	798.52	811.05
1080	763.68	776.50	789.33	802.16	814.99	827.81
1100	777.76	791.17	804.57	817.97	831.37	844.77
1120	791.62	805.53	819.44	833.36	847.27	861.18

Source: Maxwell, 1950.

Table C.3: Hydrocarbon vapour enthalpy data for MEABP = 300°F and K = 11 – 12

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	175.25	172.59	169.93	167.28	164.62	161.96
20	180.93	178.59	176.25	173.90	171.56	169.22
40	186.86	184.86	182.86	180.86	178.86	176.86
60	192.94	191.25	189.56	187.88	186.19	184.51
80	199.18	197.80	196.42	195.04	193.66	192.29
100	205.84	204.73	203.62	202.51	201.41	200.30
120	213.09	212.06	211.04	210.01	208.98	207.96
140	220.02	219.20	218.39	217.57	216.75	215.93
160	227.41	226.82	226.24	225.66	225.08	224.50
180	235.52	235.07	234.62	234.17	233.71	233.26
200	243.15	242.92	242.70	242.48	242.26	242.03
220	250.07	250.32	250.57	250.82	251.07	251.32
240	258.30	258.73	259.17	259.60	260.04	260.47
260	266.78	267.32	267.86	268.40	268.94	269.48
280	275.06	275.86	276.66	277.46	278.26	279.06
300	284.19	285.19	286.20	287.20	288.20	289.20
320	292.73	293.93	295.12	296.31	297.50	298.69
340	302.24	303.59	304.94	306.30	307.65	309.00
360	312.58	314.02	315.46	316.90	318.34	319.77
380	322.63	324.18	325.73	327.27	328.82	330.37
400	332.11	333.90	335.68	337.46	339.24	341.03
420	341.54	343.57	345.59	347.61	349.63	351.65
440	351.73	353.86	355.99	358.11	360.24	362.37
460	362.34	364.62	366.89	369.17	371.44	373.72
480	373.42	375.94	378.47	380.99	383.52	386.04
500	384.62	387.29	389.97	392.65	395.32	398.00
520	395.29	398.41	401.53	404.65	407.77	410.89
540	406.19	409.58	412.97	416.36	419.75	423.14
560	416.87	420.52	424.17	427.82	431.47	435.12
580	428.01	431.96	435.91	439.86	443.81	447.76
600	439.69	443.76	447.83	451.90	455.97	460.04
620	451.80	456.19	460.57	464.96	469.35	473.74
640	463.34	468.06	472.77	477.49	482.21	486.93
660	474.93	479.97	485.02	490.06	495.10	500.14
680	487.16	492.61	498.05	503.50	508.95	514.39
700	499.64	505.47	511.30	517.13	522.96	528.79
720	512.30	518.29	524.29	530.29	536.29	542.29
740	524.92	531.22	537.52	543.82	550.12	556.42
760	537.61	544.28	550.96	557.63	564.31	570.98
780	550.17	557.13	564.09	571.05	578.01	584.96
800	562.83	570.00	577.18	584.35	591.52	598.69
820	576.66	584.05	591.43	598.81	606.19	613.57
840	589.01	596.92	604.84	612.76	620.67	628.59
860	601.61	609.89	618.16	626.44	634.71	642.99
880	615.56	623.98	632.41	640.83	649.25	657.68
900	628.88	637.84	646.80	655.76	664.72	673.68
920	641.84	651.41	660.98	670.55	680.12	689.70
940	655.92	665.68	675.43	685.19	694.94	704.70
960	669.92	679.95	689.98	700.02	710.05	720.08
980	683.56	694.03	704.50	714.97	725.43	735.90
1000	697.91	708.57	719.23	729.89	740.55	751.21
1020	711.62	722.83	734.04	745.26	756.47	767.68
1040	725.70	737.23	748.76	760.29	771.82	783.35
1060	740.84	752.50	764.16	775.82	787.48	799.13
1080	755.66	767.71	779.75	791.80	803.85	815.89
1100	769.42	782.09	794.76	807.44	820.11	832.78
1120	784.49	797.57	810.64	823.71	836.78	849.85
1140	798.91	812.31	825.71	839.11	852.51	865.91

Source: Maxwell, 1950.

Table C. 4: Hydrocarbon liquid enthalpy data for MEABP = 300°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0.20	0.24	0.29	0.33	0.37	0.41
20	7.75	8.17	8.60	9.03	9.45	9.88
40	16.77	17.34	17.91	18.48	19.05	19.62
60	25.55	26.36	27.16	27.97	28.77	29.57
80	34.35	35.31	36.28	37.24	38.20	39.17
100	43.94	44.85	45.76	46.67	47.58	48.48
120	52.97	54.19	55.40	56.62	57.83	59.05
140	62.57	64.05	65.52	67.00	68.47	69.94
160	72.21	73.96	75.71	77.46	79.21	80.96
180	81.92	83.86	85.80	87.73	89.67	91.61
200	92.43	94.35	96.27	98.19	100.11	102.03
220	102.56	104.89	107.23	109.56	111.89	114.23
240	113.48	115.99	118.49	121.00	123.50	126.01
260	124.52	127.09	129.65	132.21	134.77	137.33
280	135.07	137.91	140.75	143.59	146.42	149.26
300	146.03	149.19	152.36	155.52	158.68	161.85
320	157.82	161.19	164.55	167.92	171.28	174.65
340	169.58	173.29	177.00	180.71	184.42	188.13
360	181.80	185.77	189.75	193.72	197.69	201.67
380	194.68	198.70	202.72	206.75	210.77	214.79
400	207.56	211.67	215.77	219.87	223.98	228.08
420	220.69	224.98	229.28	233.58	237.88	242.18
440	234.11	238.57	243.04	247.50	251.96	256.43
460	246.65	251.52	256.40	261.28	266.16	271.04
480	261.11	266.38	271.66	276.93	282.21	287.48
500	274.84	280.16	285.47	290.78	296.10	301.41
520	288.40	293.89	299.37	304.85	310.33	315.81
540	304.29	310.08	315.88	321.67	327.46	333.25
560	319.00	325.04	331.08	337.12	343.16	349.20
580	335.13	341.51	347.89	354.27	360.66	367.04
600	350.68	357.81	364.93	372.06	379.19	386.31

Source: Maxwell, 1950.

Table C.5: Hydrocarbon vapor enthalpy data for MEABP = 400°F and K = 11 – 12

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	167.62	165.22	162.82	160.42	158.02	155.62
20	174.02	171.96	169.91	167.86	165.81	163.76
40	179.94	178.13	176.32	174.51	172.69	170.88
60	186.39	184.72	183.06	181.40	179.74	178.08
80	192.50	191.12	189.73	188.35	186.96	185.58
100	199.48	198.21	196.95	195.69	194.42	193.16
120	206.21	205.32	204.44	203.55	202.66	201.78
140	213.27	212.35	211.44	210.53	209.61	208.70
160	221.95	220.89	219.83	218.77	217.71	216.65
180	228.82	228.13	227.43	226.74	226.04	225.35
200	236.82	236.41	236.00	235.58	235.17	234.76
220	244.23	243.93	243.62	243.32	243.01	242.71
240	252.48	252.21	251.93	251.65	251.37	251.10
260	261.24	261.07	260.90	260.73	260.57	260.40
280	270.09	270.06	270.02	269.99	269.95	269.92
300	278.61	278.96	279.30	279.65	279.99	280.33
320	288.11	288.36	288.61	288.85	289.10	289.35
340	297.17	297.81	298.44	299.07	299.71	300.34
360	306.84	307.54	308.24	308.95	309.65	310.36
380	316.15	316.96	317.76	318.57	319.37	320.18
400	325.50	326.72	327.93	329.15	330.37	331.58
420	336.12	337.18	338.23	339.28	340.34	341.39
440	347.06	348.41	349.77	351.12	352.47	353.82
460	356.74	358.66	360.58	362.50	364.42	366.35
480	367.24	369.30	371.35	373.41	375.46	377.52
500	378.94	380.98	383.02	385.05	387.09	389.13
520	389.90	392.33	394.75	397.17	399.59	402.02
540	400.52	403.20	405.88	408.57	411.25	413.93
560	410.40	413.67	416.95	420.22	423.50	426.77
580	422.91	426.22	429.53	432.84	436.15	439.47
600	433.45	437.00	440.55	444.10	447.65	451.20
620	445.20	448.77	452.34	455.91	459.47	463.04
640	457.32	461.21	465.09	468.98	472.86	476.75
660	469.16	473.55	477.95	482.34	486.74	491.14
680	481.18	486.07	490.96	495.85	500.74	505.63
700	492.79	497.98	503.17	508.36	513.54	518.73
720	506.17	511.35	516.53	521.71	526.89	532.06
740	517.71	523.50	529.30	535.09	540.89	546.69
760	530.52	536.53	542.54	548.55	554.56	560.57
780	543.76	549.79	555.82	561.85	567.88	573.92
800	558.15	564.11	570.07	576.04	582.00	587.96
820	570.49	576.97	583.46	589.94	596.43	602.92
840	583.01	589.92	596.82	603.72	610.62	617.53
860	596.52	603.73	610.93	618.14	625.35	632.55
880	610.01	617.44	624.87	632.30	639.74	647.17
900	622.48	630.53	638.57	646.61	654.65	662.69
920	636.94	645.05	653.15	661.26	669.36	677.46
940	649.30	657.97	666.64	675.31	683.97	692.64
960	662.75	671.74	680.73	689.72	698.72	707.71
980	677.17	686.61	696.05	705.49	714.92	724.36
1000	690.46	700.27	710.08	719.88	729.69	739.50
1020	705.18	715.34	725.51	735.67	745.84	756.00
1040	720.34	730.57	740.79	751.01	761.24	771.46
1060	735.34	745.63	755.93	766.23	776.53	786.83
1080	749.35	760.34	771.34	782.33	793.32	804.32
1100	763.82	775.04	786.25	797.47	808.69	819.90
1120	778.70	790.46	802.22	813.99	825.75	837.52
1140	793.54	805.70	817.85	830.00	842.15	854.30

Source: Maxwell, 1950.

Table C.6: Hydrocarbon liquid enthalpy data for MEABP = 400°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.70	7.90	8.10	8.30	8.50	8.70
40	15.46	15.93	16.39	16.86	17.32	17.79
60	24.16	24.88	25.59	26.31	27.02	27.74
80	31.58	32.54	33.50	34.46	35.42	36.39
100	39.75	40.86	41.97	43.08	44.18	45.29
120	49.31	50.57	51.83	53.09	54.35	55.62
140	58.42	59.89	61.36	62.84	64.31	65.78
160	67.37	69.02	70.67	72.31	73.96	75.60
180	77.24	78.95	80.65	82.36	84.07	85.78
200	86.99	88.87	90.75	92.62	94.50	96.38
220	96.32	98.37	100.41	102.46	104.51	106.56
240	107.95	109.93	111.91	113.89	115.86	117.84
260	118.58	120.66	122.74	124.82	126.91	128.99
280	129.36	131.66	133.96	136.26	138.56	140.86
300	140.21	142.89	145.56	148.23	150.91	153.58
320	151.32	154.05	156.78	159.51	162.24	164.97
340	163.26	166.17	169.08	171.99	174.90	177.81
360	175.32	178.44	181.56	184.68	187.80	190.92
380	187.16	190.23	193.30	196.37	199.45	202.52
400	200.18	203.39	206.60	209.82	213.03	216.25
420	213.59	216.77	219.95	223.13	226.31	229.50
440	225.94	229.45	232.95	236.46	239.97	243.48
460	238.23	242.03	245.83	249.63	253.43	257.23
480	251.00	255.12	259.23	263.35	267.46	271.57
500	264.20	268.52	272.83	277.15	281.47	285.79
520	277.65	282.44	287.23	292.02	296.81	301.60
540	291.31	296.15	300.98	305.81	310.65	315.48
560	305.20	310.51	315.82	321.13	326.44	331.75
580	320.08	325.59	331.10	336.62	342.13	347.64
600	334.63	340.57	346.51	352.45	358.38	364.32
620	349.72	356.08	362.45	368.81	375.17	381.53
640	365.41	372.19	378.96	385.73	392.50	399.27
660	380.81	387.91	395.02	402.12	409.22	416.32
680	397.78	404.95	412.13	419.31	426.48	433.66

Source: Maxwell, 1950.

Table C. 7: Hydrocarbon vapor enthalpy data for MEABP = 500°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	165.62	162.14	158.65	155.17	151.69	148.20
20	171.50	168.21	164.92	161.62	158.33	155.04
40	177.35	174.38	171.42	168.45	165.49	162.52
60	184.01	181.21	178.40	175.60	172.79	169.98
80	190.03	187.57	185.10	182.63	180.17	177.70
100	197.11	194.84	192.58	190.31	188.04	185.77
120	203.34	201.23	199.12	197.01	194.90	192.79
140	210.66	208.55	206.45	204.34	202.23	200.13
160	217.84	216.02	214.20	212.39	210.57	208.76
180	225.78	224.08	222.39	220.69	218.99	217.30
200	233.16	231.55	229.94	228.33	226.71	225.10
220	241.18	239.96	238.73	237.51	236.29	235.06
240	249.49	248.36	247.22	246.09	244.96	243.82
260	257.24	256.31	255.39	254.46	253.54	252.61
280	265.65	265.00	264.34	263.69	263.03	262.38
300	275.16	274.42	273.67	272.93	272.19	271.45
320	285.10	284.30	283.50	282.70	281.90	281.11
340	294.83	293.99	293.16	292.32	291.49	290.65
360	304.42	303.73	303.04	302.35	301.65	300.96
380	313.71	313.39	313.08	312.77	312.46	312.15
400	323.59	323.49	323.39	323.30	323.20	323.11
420	332.69	333.01	333.32	333.64	333.95	334.26
440	342.76	343.11	343.46	343.81	344.16	344.50
460	352.66	353.38	354.09	354.80	355.52	356.23
480	364.52	365.34	366.17	366.99	367.82	368.64
500	375.06	376.31	377.57	378.82	380.08	381.33
520	385.70	387.30	388.90	390.50	392.10	393.70
540	397.51	399.10	400.70	402.30	403.89	405.49
560	409.14	410.90	412.66	414.42	416.18	417.94
580	419.94	421.95	423.96	425.96	427.97	429.98
600	431.32	433.56	435.80	438.05	440.29	442.53
620	443.43	445.72	448.01	450.30	452.59	454.88
640	455.04	457.57	460.09	462.62	465.14	467.67
660	467.10	469.93	472.76	475.59	478.42	481.25
680	477.88	481.32	484.75	488.19	491.62	495.05
700	490.28	493.91	497.55	501.18	504.81	508.45
720	502.85	506.65	510.45	514.24	518.04	521.83
740	514.89	519.07	523.24	527.42	531.60	535.77
760	527.83	532.18	536.53	540.87	545.22	549.57
780	539.95	544.71	549.47	554.23	558.99	563.75
800	553.32	558.47	563.62	568.77	573.91	579.06
820	566.16	571.68	577.19	582.71	588.22	593.73
840	578.50	584.25	590.00	595.75	601.50	607.25
860	592.13	597.94	603.74	609.55	615.35	621.16
880	606.12	612.25	618.39	624.52	630.66	636.79
900	618.17	625.11	632.04	638.98	645.91	652.85
920	632.49	639.38	646.27	653.16	660.05	666.94
940	645.97	653.16	660.35	667.54	674.72	681.91
960	660.26	667.76	675.27	682.77	690.27	697.77
980	675.02	682.68	690.34	698.00	705.66	713.32
1000	689.02	696.98	704.95	712.91	720.88	728.84
1020	704.02	712.26	720.50	728.74	736.97	745.21
1040	718.15	726.74	735.33	743.93	752.52	761.11
1060	731.92	741.08	750.24	759.41	768.57	777.74
1080	748.06	757.47	766.88	776.28	785.69	795.09
1100	762.00	772.00	781.99	791.98	801.98	811.97
1120	775.97	786.43	796.89	807.35	817.81	828.27
1140	790.71	801.52	812.34	823.15	833.97	844.78
1160	806.26	817.36	828.45	839.55	850.65	861.74

Source: Maxwell, 1950.

Table C.8: Hydrocarbon liquid enthalpy data for MEABP = 500°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	8.34	8.55	8.76	8.97	9.18	9.40
40	16.33	16.42	16.51	16.60	16.70	16.79
60	24.17	24.34	24.51	24.69	24.86	25.03
80	32.51	32.93	33.34	33.76	34.18	34.59
100	40.09	40.74	41.40	42.06	42.71	43.37
120	48.80	49.68	50.56	51.44	52.33	53.21
140	58.68	59.62	60.57	61.51	62.46	63.41
160	67.31	68.39	69.48	70.57	71.65	72.74
180	77.26	78.24	79.22	80.20	81.18	82.16
200	86.56	87.86	89.15	90.45	91.75	93.04
220	95.50	97.15	98.80	100.45	102.10	103.76
240	106.01	107.79	109.58	111.36	113.15	114.93
260	117.04	118.85	120.65	122.45	124.25	126.05
280	127.04	129.10	131.16	133.22	135.28	137.33
300	137.82	140.09	142.36	144.63	146.90	149.17
320	149.33	151.47	153.61	155.74	157.88	160.02
340	159.02	161.68	164.34	167.00	169.66	172.32
360	171.17	173.89	176.62	179.35	182.07	184.80
380	182.97	185.66	188.34	191.03	193.72	196.40
400	194.53	197.33	200.13	202.93	205.73	208.53
420	205.53	208.75	211.97	215.19	218.41	221.63
440	218.36	221.56	224.75	227.94	231.14	234.33
460	230.10	233.59	237.09	240.58	244.08	247.57
480	242.83	246.37	249.92	253.47	257.02	260.57
500	255.38	259.05	262.73	266.40	270.07	273.74
520	268.39	272.46	276.54	280.62	284.70	288.77
540	283.06	287.09	291.11	295.14	299.17	303.19
560	295.46	300.25	305.04	309.82	314.61	319.40
580	310.38	315.43	320.48	325.53	330.59	335.64
600	325.04	330.17	335.31	340.45	345.59	350.73
620	339.59	345.05	350.51	355.96	361.42	366.88
640	353.23	358.89	364.56	370.22	375.88	381.55
660	367.88	373.80	379.71	385.62	391.54	397.45
680	383.10	389.41	395.73	402.04	408.36	414.68
700	398.75	405.14	411.53	417.93	424.32	430.71
720	413.92	420.81	427.69	434.58	441.46	448.35
740	430.95	438.00	445.05	452.10	459.15	466.20
760	446.71	454.06	461.41	468.76	476.11	483.46
780	463.72	471.16	478.61	486.06	493.51	500.95
800	479.61	487.70	495.78	503.87	511.96	520.05

Source: Maxwell, 1950.

Table C.9: Hydrocarbon liquid enthalpy data for MEABP = 600 ^{o}F and K = 11 – 12.

T ($^{\circ}$F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.92	8.06	8.20	8.33	8.47	8.61
40	15.09	15.52	15.94	16.36	16.78	17.20
60	23.22	23.76	24.30	24.85	25.39	25.93
80	32.36	32.87	33.39	33.91	34.42	34.94
100	39.91	40.75	41.59	42.43	43.27	44.11
120	48.39	49.37	50.34	51.32	52.30	53.27
140	56.84	57.94	59.05	60.15	61.25	62.36
160	65.97	67.13	68.29	69.45	70.61	71.76
180	75.95	77.08	78.20	79.33	80.45	81.57
200	85.29	86.55	87.80	89.05	90.31	91.56
220	94.60	96.04	97.48	98.92	100.36	101.80
240	104.68	106.13	107.58	109.03	110.48	111.93
260	114.30	115.96	117.62	119.28	120.94	122.61
280	123.68	125.66	127.65	129.63	131.62	133.61
300	134.60	136.55	138.51	140.46	142.42	144.37
320	145.15	147.36	149.56	151.76	153.96	156.16
340	156.17	158.55	160.93	163.31	165.69	168.07
360	167.38	169.90	172.42	174.94	177.47	179.99
380	178.42	181.16	183.89	186.63	189.37	192.11
400	189.95	192.82	195.69	198.56	201.43	204.30
420	201.97	204.94	207.91	210.88	213.85	216.82
440	214.53	217.58	220.64	223.69	226.75	229.80
460	226.92	230.02	233.13	236.23	239.34	242.44
480	239.57	242.96	246.36	249.76	253.15	256.55
500	252.55	255.81	259.06	262.32	265.58	268.84
520	264.31	267.69	271.07	274.44	277.82	281.20
540	277.65	281.20	284.75	288.30	291.85	295.40
560	290.79	294.62	298.46	302.29	306.13	309.96
580	303.45	307.57	311.68	315.80	319.92	324.03
600	317.22	321.17	325.11	329.05	333.00	336.94
620	331.49	335.58	339.67	343.76	347.85	351.95
640	345.88	350.23	354.59	358.94	363.30	367.65
660	359.94	364.60	369.25	373.90	378.56	383.21
680	374.04	379.04	384.04	389.04	394.04	399.04
700	388.85	394.13	399.42	404.70	409.98	415.26
720	403.65	409.19	414.74	420.28	425.83	431.37
740	419.07	424.95	430.83	436.72	442.60	448.48
760	434.54	440.80	447.07	453.34	459.61	465.88
780	449.97	456.58	463.19	469.79	476.40	483.00
800	466.08	472.93	479.78	486.63	493.48	500.33
820	481.86	489.12	496.37	503.63	510.88	518.14
840	497.97	505.71	513.46	521.20	528.95	536.69
860	515.04	523.28	531.53	539.77	548.01	556.26
880	530.48	539.65	548.81	557.97	567.14	576.30
900	548.65	558.02	567.40	576.77	586.15	595.52

Source: Maxwell, 1950.

Table C.10: Hydrocarbon vapor enthalpy data for MEABP = 600°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	159.95	156.45	152.95	149.45	145.95	142.45
20	166.78	163.39	160.00	156.61	153.22	149.83
40	172.72	169.52	166.33	163.13	159.93	156.73
60	178.50	175.62	172.73	169.84	166.96	164.07
80	184.91	182.23	179.55	176.87	174.19	171.51
100	192.18	189.47	186.76	184.05	181.34	178.63
120	199.43	196.84	194.24	191.65	189.06	186.47
140	206.08	203.70	201.31	198.93	196.55	194.17
160	213.33	211.14	208.95	206.75	204.56	202.37
180	221.11	219.07	217.04	215.00	212.96	210.93
200	228.41	226.52	224.63	222.74	220.85	218.96
220	236.70	234.92	233.14	231.37	229.59	227.81
240	244.40	242.89	241.38	239.87	238.36	236.85
260	252.48	251.10	249.72	248.34	246.96	245.58
280	261.25	259.93	258.62	257.31	256.00	254.69
300	269.74	268.74	267.74	266.74	265.74	264.74
320	279.44	278.43	277.42	276.41	275.40	274.39
340	288.20	287.18	286.16	285.14	284.12	283.11
360	297.15	296.45	295.75	295.05	294.35	293.65
380	306.77	306.44	306.11	305.78	305.45	305.12
400	315.66	315.65	315.63	315.61	315.60	315.58
420	326.10	326.28	326.47	326.65	326.84	327.02
440	336.49	336.54	336.60	336.65	336.70	336.76
460	346.54	346.85	347.15	347.46	347.77	348.07
480	357.46	358.19	358.93	359.67	360.40	361.14
500	369.03	369.66	370.28	370.91	371.53	372.16
520	380.07	380.92	381.78	382.63	383.49	384.34
540	390.63	391.82	393.00	394.19	395.38	396.56
560	401.89	403.19	404.49	405.79	407.09	408.39
580	413.01	414.50	416.00	417.50	418.99	420.49
600	423.35	425.15	426.95	428.75	430.54	432.34
620	436.02	437.74	439.47	441.19	442.92	444.65
640	448.66	450.51	452.35	454.20	456.05	457.90
660	459.97	462.29	464.61	466.94	469.26	471.58
680	471.32	474.02	476.72	479.43	482.13	484.83
700	484.30	487.00	489.70	492.40	495.09	497.79
720	497.51	500.36	503.20	506.05	508.89	511.73
740	509.77	513.11	516.44	519.78	523.12	526.46
760	522.35	526.05	529.75	533.44	537.14	540.84
780	535.36	539.25	543.15	547.05	550.94	554.84
800	547.58	551.93	556.29	560.64	565.00	569.35
820	561.05	565.64	570.23	574.83	579.42	584.01
840	574.27	578.95	583.62	588.30	592.97	597.65
860	587.07	592.07	597.07	602.07	607.06	612.06
880	600.32	605.69	611.07	616.45	621.82	627.20
900	613.87	619.47	625.08	630.68	636.29	641.89
920	626.79	632.93	639.07	645.20	651.34	657.48
940	640.17	646.58	652.99	659.40	665.80	672.21
960	654.83	661.39	667.96	674.52	681.09	687.65
980	669.29	676.16	683.02	689.89	696.75	703.62
1000	682.24	689.52	696.80	704.08	711.36	718.64
1020	696.85	704.36	711.87	719.37	726.88	734.39
1040	711.12	718.94	726.75	734.56	742.38	750.19
1060	725.03	733.23	741.43	749.63	757.83	766.03
1080	739.54	748.10	756.65	765.21	773.76	782.31
1100	753.82	762.86	771.90	780.93	789.97	799.00
1120	768.44	777.86	787.28	796.71	806.13	815.55
1140	783.69	793.25	802.80	812.36	821.91	831.47

Source: Maxwell, 1950.

Table C.11: Hydrocarbon liquid enthalpy data for MEABP = 800°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	0	0	0	0	0	0
20	7.74	7.81	7.87	7.94	8.00	8.07
40	15.25	15.41	15.56	15.72	15.87	16.02
60	22.46	22.76	23.06	23.35	23.65	23.95
80	30.81	31.15	31.48	31.82	32.16	32.50
100	38.70	39.22	39.73	40.25	40.76	41.27
120	46.46	47.15	47.84	48.53	49.22	49.91
140	54.20	55.07	55.95	56.82	57.69	58.57
160	63.97	64.75	65.53	66.31	67.08	67.86
180	72.44	73.29	74.13	74.98	75.83	76.67
200	81.27	82.28	83.30	84.31	85.32	86.34
220	89.89	91.19	92.48	93.78	95.08	96.38
240	100.40	101.50	102.59	103.69	104.78	105.88
260	110.69	111.64	112.58	113.52	114.46	115.41
280	120.32	121.50	122.68	123.87	125.05	126.23
300	130.81	132.05	133.29	134.53	135.77	137.01
320	140.10	141.58	143.05	144.52	145.99	147.47
340	151.40	152.88	154.36	155.84	157.32	158.80
360	161.59	163.42	165.25	167.09	168.92	170.75
380	172.03	173.82	175.61	177.40	179.20	180.99
400	183.51	185.38	187.24	189.10	190.97	192.83
420	195.52	197.35	199.17	201.00	202.83	204.65
440	205.96	208.14	210.33	212.51	214.70	216.89
460	218.32	220.59	222.85	225.12	227.38	229.65
480	230.42	232.78	235.14	237.50	239.87	242.23
500	242.71	245.17	247.62	250.08	252.54	255.00
520	254.97	257.53	260.09	262.65	265.21	267.77
540	267.83	270.54	273.26	275.98	278.70	281.42
560	280.18	283.23	286.27	289.31	292.35	295.40
580	293.20	296.48	299.76	303.05	306.33	309.61
600	305.91	309.51	313.12	316.72	320.33	323.93
620	318.91	322.70	326.49	330.28	334.07	337.86
640	332.82	336.78	340.74	344.70	348.66	352.63
660	346.82	350.78	354.74	358.69	362.65	366.60
680	360.05	364.35	368.66	372.97	377.28	381.59
700	373.72	378.30	382.88	387.46	392.04	396.63
720	387.57	392.24	396.91	401.57	406.24	410.91
740	402.54	407.32	412.11	416.90	421.69	426.48
760	417.13	422.11	427.09	432.07	437.05	442.04
780	431.65	436.84	442.03	447.23	452.42	457.61
800	447.21	452.47	457.73	462.98	468.24	473.50
820	461.39	466.85	472.30	477.75	483.20	488.66
840	477.79	483.40	489.00	494.60	500.20	505.81
860	493.26	499.13	504.99	510.86	516.72	522.58
880	509.35	515.19	521.03	526.87	532.72	538.56
900	524.08	530.32	536.56	542.81	549.05	555.29
920	540.55	546.99	553.43	559.87	566.32	572.76
940	557.03	563.65	570.27	576.89	583.51	590.13
960	573.65	580.53	587.41	594.30	601.18	608.06
980	589.75	597.06	604.37	611.67	618.98	626.29
1000	606.56	614.20	621.83	629.47	637.11	644.75
1020	623.47	631.27	639.07	646.87	654.66	662.46
1040	641.90	650.25	658.59	666.94	675.29	683.64
1060	659.93	668.58	677.23	685.87	694.52	703.17
1080	677.10	686.52	695.95	705.38	714.81	724.23
1100	695.64	705.56	715.48	725.40	735.32	745.24

Source: Maxwell, 1950.

Table C.12:Hydrocarbon vapor enthalpy data for MEABP = 800°F and K = 11 – 12.

T (°F)	Hydrocarbon Enthalpy (Btu/lb) for various values of K					
	11	11.2	11.4	11.6	11.8	12
0	147.65	143.99	140.33	136.67	133.01	129.35
20	152.66	149.20	145.74	142.29	138.83	135.38
40	159.17	155.88	152.58	149.29	145.99	142.70
60	165.75	162.48	159.20	155.92	152.64	149.36
80	171.83	168.66	165.49	162.32	159.15	155.98
100	178.70	175.74	172.77	169.81	166.85	163.89
120	185.20	182.35	179.49	176.64	173.79	170.93
140	192.28	189.64	187.00	184.36	181.72	179.07
160	199.87	197.35	194.84	192.32	189.81	187.29
180	207.11	204.67	202.24	199.80	197.37	194.93
200	215.73	213.29	210.85	208.41	205.97	203.53
220	223.77	221.47	219.17	216.87	214.57	212.27
240	231.59	229.44	227.30	225.16	223.01	220.87
260	239.35	237.48	235.61	233.74	231.87	230.00
280	248.35	246.60	244.85	243.10	241.35	239.60
300	257.59	255.83	254.08	252.32	250.57	248.81
320	266.19	264.51	262.82	261.13	259.45	257.76
340	275.31	273.71	272.10	270.49	268.88	267.27
360	285.00	283.47	281.93	280.39	278.85	277.31
380	294.05	292.81	291.58	290.34	289.10	287.87
400	303.97	302.92	301.88	300.84	299.79	298.75
420	312.78	311.94	311.10	310.26	309.41	308.57
440	323.67	322.70	321.72	320.75	319.77	318.80
460	334.13	333.45	332.78	332.10	331.43	330.76
480	344.19	343.63	343.07	342.51	341.95	341.39
500	356.14	355.67	355.20	354.73	354.26	353.79
520	367.42	367.63	367.84	368.05	368.26	368.47
540	377.61	377.95	378.28	378.62	378.95	379.29
560	388.30	388.94	389.58	390.22	390.86	391.50
580	401.02	401.48	401.95	402.42	402.89	403.36
600	410.89	411.83	412.77	413.71	414.65	415.59
620	422.64	423.84	425.05	426.25	427.45	428.66
640	434.21	435.63	437.04	438.45	439.86	441.28
660	445.97	447.57	449.17	450.77	452.37	453.97
680	458.16	459.94	461.73	463.51	465.29	467.08
700	470.66	472.55	474.45	476.34	478.23	480.13
720	482.53	484.77	487.00	489.23	491.46	493.69
740	495.57	497.91	500.26	502.61	504.95	507.30
760	508.08	510.65	513.22	515.79	518.36	520.93
780	520.96	523.72	526.49	529.25	532.01	534.77
800	534.29	537.22	540.15	543.08	546.02	548.95
820	547.30	550.54	553.77	557.00	560.24	563.47
840	560.83	564.13	567.42	570.72	574.01	577.31
860	574.74	578.08	581.41	584.75	588.09	591.43
880	587.02	590.80	594.59	598.37	602.16	605.94
900	600.30	604.39	608.49	612.59	616.68	620.78
920	613.85	618.38	622.91	627.45	631.98	636.52
940	628.26	633.00	637.74	642.49	647.23	651.97
960	642.25	647.20	652.16	657.12	662.08	667.04
980	655.83	661.14	666.45	671.76	677.07	682.38
1000	670.38	675.96	681.54	687.13	692.71	698.29
1020	684.46	690.48	696.51	702.54	708.56	714.59
1040	698.46	704.76	711.06	717.36	723.66	729.96
1060	712.70	719.35	726.00	732.65	739.29	745.94
1080	727.97	734.95	741.94	748.92	755.91	762.89
1100	741.39	748.93	756.48	764.02	771.56	779.11
1120	756.85	764.55	772.26	779.97	787.67	795.38
1140	771.12	779.40	787.67	795.95	804.23	812.50
1160	787.50	796.00	804.50	813.00	821.50	830.01

Source: Maxwell, 1950.

Table C.13: Vapor pressure data for hydrocarbons

T (°F)	C4H10	C5H12	100	150	200	250	300	350	400	450	500	550	600	700	800	900	1000	1100	1200
50	1.47	0.38	0.36	0.12	0.04	0.10													
75	2.32	0.65	0.61	0.22	0.07	0.23	0.01												
100	3.51	1.07	1.00	0.39	0.14	0.47	0.01												
125	5.20	1.64	1.53	0.65	0.25	0.90	0.03	0.01											
150	7.25	2.46	2.32	0.99	0.42	0.16	0.06	0.02	0.01										
175	9.85	3.51	3.35	1.50	0.66	0.27	0.10	0.04	0.01										
200	13.10	4.96	4.56	2.22	0.99	0.44	0.17	0.07	0.02	0.01									
225	17.15	6.78	6.50	3.15	1.48	0.67	0.29	0.12	0.04	0.02	0.01								
250	21.95	9.01	8.56	4.34	2.11	1.01	0.45	0.19	0.08	0.03	0.01								
275	27.91	12.02	11.40	5.82	3.02	1.47	0.68	0.31	0.13	0.05	0.02	0.01							
300	34.90	15.20	14.40	7.78	4.05	2.08	1.00	0.47	0.21	0.09	0.04	0.01							
325	42.40	19.30	18.80	10.05	5.45	2.81	1.43	0.70	0.33	0.14	0.06	0.02	0.01						
350	51.80	24.30	23.20	13.10	7.41	3.80	2.02	1.02	0.48	0.23	0.10	0.04	0.02						
375	60.50	30.20	28.90	16.40	9.42	5.15	2.81	1.45	0.71	0.35	0.16	0.07	0.03						
400	72.10	37.10	35.02	20.70	12.02	6.76	3.68	1.94	1.01	0.51	0.25	0.11	0.05	0.01					
425	83.20	43.60	41.80	25.20	14.90	8.55	4.76	2.66	1.39	0.74	0.36	0.17	0.08	0.01					
450	95.30	51.20	49.10	30.80	18.30	10.70	6.22	3.48	1.86	1.01	0.51	0.25	0.12	0.02					
475		60.00	57.90	36.50	22.20	13.20	7.83	4.51	2.44	1.41	0.72	0.36	0.18	0.04	0.01				
500		70.10	67.40	43.20	27.30	16.30	9.80	5.77	3.21	1.84	1.02	0.52	0.26	0.06	0.01				
525		80.20	77.50	50.10	32.50	19.90	12.40	7.42	4.24	2.48	1.23	0.74	0.38	0.09	0.02				
550		92.00	88.70	59.10	39.10	24.10	15.20	9.23	5. 42	3.24	1.81	1.01	0.53	0.14	0.03	0.01			
575			100.00	69.20	45.20	29.20	18.60	11.60	6.97	4.20	2.42	1.35	0.74	0.20	0.05	0.01			
600				79.70	52.70	34.10	22.30	14.00	8.48	5.22	3.11	1.79	1.01	0.29	0.08	0.02			
650				100.00	71.20	47.60	31.60	20.60	13.00	8.20	5.04	3.05	1.72	0.56	0.17	0.05	0.01		
700					90.00	62.20	42.10	28.40	18.70	12.20	7.65	4.68	2.83	0.98	0.33	0.10	0.03	0.01	
750						79.10	55.20	38.10	25.90	17.60	11.20	7.15	4.24	1.62	0.61	0.20	0.06	0.01	
800						96.70	70.00	49.70	33.50	23.50	15.30	10.70	6.48	2.52	1.02	0.37	0.11	0.03	0.01
850							86.30	63.80	44.10	31.80	20.80	14.80	9.25	3.76	1.61	0.62	0.20	0.06	0.02
900								78.50	57.20	41.70	28.10	19.10	12.60	5.58	2.43	1.01	0.36	0.12	0.03
950								96.10	71.90	53.50	37.10	26.20	17.90	8.05	3.69	1.59	0.62	0.22	0.07
1000									87.40	68.10	47.90	34.80	23.20	11.30	5.36	2.45	1.02	0.39	0.13
1050										83.90	59.60	44.50	30.80	15.20	7.70	3.63	1.59	0.64	0.24
1100										100.00	73.80	56.10	39.10	20.40	10.80	5.22	2.32	0.99	0.42
1150											89.90	68.50	50.20	26.30	14.10	7.36	3.43	1.52	0.65
1200												82.00	62.40	32.40	18.40	9.90	4.86	2.21	1.01

Source: Uppaluri 2010.

Table C.14: EFV-TBP correlation data presented by Maxwell (1950).

Maxwell's correlation 1		Maxwell's correlation 2 T_{50} (DRL) < 300 °F		Maxwell's correlation 2 T_{50} (DRL) > 300 °F		Maxwell's correlation 3	
S_{TBP}	S_{FRL}	S_{TBP}	ΔT_{50} (DRL – FRL)	S_{TBP}	ΔT_{50} (DRL – FRL)	Percent off	$\frac{\Delta T50\ (Flash - FRL)}{\Delta T50\ (DRL - FRL)}$
0.02	7.74	0.02	7.74	2.00	0.65	1.08	0.33
0.48	7.71	0.48	7.71	2.46	3.33	3.49	0.37
0.88	7.69	0.88	7.69	2.84	5.62	7.49	0.39
1.34	7.27	1.34	7.27	3.09	8.32	13.47	0.38
1.76	8.02	1.76	8.02	3.39	11.39	18.66	0.35
2.16	9.15	2.16	9.15	3.65	14.47	24.64	0.34
2.58	11.06	2.58	11.06	3.91	19.10	30.63	0.34
3.00	12.19	3.00	12.19	4.17	23.34	35.23	0.34
3.31	14.11	3.31	14.11	4.61	28.73	40.02	0.34
3.67	17.96	3.67	17.96	4.97	32.97	45.21	0.34
3.97	22.20	3.97	22.20	5.25	34.50	50.80	0.34
4.37	27.20	4.37	27.20	5.53	36.03	54.19	0.34
4.85	31.82	4.85	31.82	5.79	36.40	60.19	0.34
5.23	34.50	5.23	34.50	6.13	37.93	66.57	0.34
5.69	36.79	5.69	36.79	6.39	38.30	71.17	0.34
6.19	37.92	6.19	37.92	6.61	39.06	75.36	0.34
6.91	39.04	6.91	39.04	6.95	39.42	79.35	0.35
7.43	39.39	7.43	39.39	7.21	39.41	83.75	0.34
7.87	38.98	7.87	38.98	7.55	39.77	87.54	0.34
8.39	39.33	8.39	39.33	8.05	39.74	91.13	0.34
8.95	39.68	8.95	39.68	8.63	39.31	94.93	0.34
9.54	39.64	9.54	39.64	9.26	40.05	98.92	0.34
10.14	39.60	10.14	39.60	9.78	39.63		
10.74	39.95	10.74	39.95	10.40	39.97		
11.24	39.53	11.24	39.53	10.88	39.94		
11.68	39.89	11.68	39.89	11.44	39.91		
11.98	39.87	11.98	39.87	11.78	39.50		
9.60	6.45			11.78	39.50		
9.88	6.71						
10.18	7.01						

Source: Maxwell, 1950.

Table C.15 :Variation of specific gravity with temperature (a) Data range: SG = 0.5 to 0.7 at 60 ^{o}F (b) Data range: SG = 0.72 to 0.98 at 60 ^{o}F. Source: Uppalluri 2010

(a)

SG = 0.5		SG = 0.52		SG = 0.54		SG = 0.58		SG = 0.6		SG = 0.62		SG = 0.64		SG = 0.66		SG = 0.68		SG = 0.7	
T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG
80.07	0.48	81.38	0.50	79.02	0.52	80.39	0.56	87.81	0.58	86.67	0.60	90.42	0.62	85.62	0.64	101.59	0.65	98.01	0.68
100.84	0.46	102.17	0.48	101.03	0.51	99.96	0.55	115.94	0.56	117.26	0.58	122.24	0.60	110.09	0.63	139.54	0.64	139.62	0.66
119.16	0.44	121.73	0.47	120.59	0.49	121.98	0.53	137.96	0.54	146.61	0.56	151.60	0.58	138.24	0.62	171.36	0.62	181.23	0.63
139.92	0.42	141.28	0.45	140.15	0.48	141.55	0.52	164.86	0.52	177.18	0.54	182.18	0.56	168.82	0.60	201.95	0.60	217.94	0.61
160.66	0.39	160.82	0.43	160.93	0.46	163.55	0.50	197.88	0.50	204.10	0.52	217.67	0.54	200.63	0.58	233.77	0.58	257.10	0.59
178.94	0.36	184.00	0.40	180.47	0.44	183.12	0.49	227.23	0.48	228.56	0.51	251.93	0.52	230.00	0.56	260.68	0.57	287.69	0.58
182.58	0.35	198.64	0.38	213.42	0.39	200.23	0.47	262.66	0.45	267.70	0.48	293.50	0.49	259.37	0.55	288.83	0.55	339.09	0.55
		212.02	0.35	232.93	0.37	216.12	0.46	289.53	0.42	301.92	0.45	330.20	0.47	285.06	0.53	335.32	0.53	377.03	0.53
				239.00	0.35	249.10	0.43	309.05	0.39	331.24	0.42	368.10	0.44	322.99	0.51	367.14	0.51	411.29	0.51
						267.40	0.40	323.68	0.37	367.87	0.38	397.43	0.41	351.12	0.49	395.26	0.49	445.55	0.49
						289.37	0.37			386.16	0.35	427.95	0.38	373.13	0.47	430.74	0.47	487.14	0.46
						301.54	0.35					446.23	0.35	403.72	0.45	463.76	0.44	520.17	0.44
														428.16	0.43	491.89	0.42	549.52	0.42
														453.82	0.41	523.66	0.39	580.08	0.39
														483.13	0.38	557.85	0.36	604.51	0.37
														501.44	0.35			616.69	0.35

(b)

SG = 0.72		SG = 0.74		SG = 0.76		SG = 0.78		SG = 0.8		SG = 0.82		SG = 0.84		SG = 0.88		SG = 0.94		SG = 0.98	
T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG	T (°F)	SG
100.53	0.70	114.07	0.71	101.91	0.74	100.78	0.76	102.08	0.78	118.07	0.79	101.01	0.82	119.54	0.86	111.21	0.92	104.01	0.96
140.91	0.67	165.48	0.68	139.85	0.72	142.41	0.74	160.85	0.76	175.62	0.77	157.35	0.80	185.67	0.83	178.55	0.90	154.23	0.95
181.30	0.65	225.45	0.66	182.70	0.70	188.92	0.72	223.29	0.73	230.73	0.75	208.78	0.78	248.12	0.81	254.48	0.87	195.86	0.93
221.69	0.63	285.43	0.63	224.33	0.68	237.88	0.70	283.28	0.71	291.96	0.73	263.89	0.77	311.80	0.79	310.81	0.85	244.85	0.92
263.30	0.61	341.73	0.60	263.49	0.66	286.87	0.68	344.51	0.68	343.38	0.71	311.63	0.74	375.46	0.76	385.51	0.82	302.41	0.90
293.90	0.60	401.70	0.57	299.00	0.65	323.58	0.66	404.50	0.66	398.48	0.69	353.27	0.73	451.38	0.73	440.62	0.80	357.52	0.88
336.73	0.57	462.88	0.53	336.94	0.63	361.54	0.65	458.35	0.63	462.13	0.66	404.69	0.71	515.04	0.70	485.92	0.79	413.85	0.86
383.24	0.55	522.83	0.50	381.00	0.61	404.38	0.63	504.86	0.61	541.70	0.62	447.54	0.69	581.14	0.67	544.70	0.76	477.53	0.84
424.83	0.52	569.31	0.47	422.62	0.59	434.97	0.61	550.15	0.58	593.09	0.59	491.60	0.67	655.81	0.64	588.77	0.74	524.07	0.82
477.45	0.50	612.12	0.44	465.46	0.56	475.36	0.59	588.09	0.57	635.92	0.57	530.78	0.65	721.89	0.60	634.06	0.72	576.72	0.80
520.27	0.47	659.79	0.40	503.38	0.54	524.32	0.56	635.80	0.54	675.08	0.55	566.26	0.63	773.28	0.57	690.39	0.70	624.47	0.78
565.53	0.44			549.88	0.51	564.69	0.54	671.28	0.52	721.55	0.51	604.22	0.61	816.10	0.54	752.83	0.68	668.54	0.76
610.77	0.40			593.90	0.48	603.84	0.52	710.43	0.49	772.92	0.48	645.83	0.59	879.71	0.51	810.37	0.65	717.53	0.74
				639.17	0.46	644.21	0.49	744.66	0.46	814.50	0.45	686.20	0.57			865.44	0.62	766.51	0.72
				673.41	0.43	678.45	0.47	776.44	0.44	842.62	0.43	722.89	0.54			903.38	0.60	815.47	0.70
				702.73	0.40	712.68	0.44	811.86	0.40			754.71	0.52					863.23	0.68
						756.68	0.40					785.30	0.51					904.86	0.67
												819.54	0.48						
												850.10	0.46						

TableC.16: Fractionation criteria correlation data for naphtha-kerosene products.

$\Delta TBP_{50\%}$ = 100 °F		$\Delta TBP_{50\%}$ = 150 °F		$\Delta TBP_{50\%}$ = 200 °F		$\Delta TBP_{50\%}$ = 250 °F	
ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N
23.30	100.63	45.40	101.75	65.18	101.56	81.08	102.75
22.13	69.63	43.85	70.40	64.79	81.21	79.92	49.84
19.81	50.12	41.52	50.02	63.63	60.80	77.59	40.39
15.54	34.69	38.03	34.62	62.85	49.92	76.81	34.04
12.83	29.24	33.38	24.60	61.30	39.92	75.26	29.85
9.72	24.65	29.50	19.93	59.75	34.54	73.32	24.50
6.62	21.62	22.13	14.93	56.26	24.54	69.83	19.60
4.29	19.46	14.76	11.64	53.16	19.37	64.79	14.68
-0.36	16.41	9.72	10.08	46.95	14.90	59.75	12.06
-3.85	14.78	5.46	8.84	39.97	11.02	54.32	9.90
-7.73	13.48	0.80	7.96	36.09	9.92	50.83	8.80
-11.22	12.46	-3.85	7.07	29.50	8.04	46.95	7.92
-15.87	10.93					45.01	7.32
-19.36	9.84						
-24.79	8.98						
-28.67	8.19						

Source: Uppaluri 2010

Table C.17: Fractionation criteria correlation data for side stream-side stream products.

$\Delta TBP_{50\%}$ = 100 °F		$\Delta TBP_{50\%}$ = 150 °F		$\Delta TBP_{50\%}$ = 200 °F		$\Delta TBP_{50\%}$ = 250 °F		$\Delta TBP_{50\%}$ = 300 °F	
ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N	ASTM Gap (°F)	F = RR x N
-24.82	5.72	-16.37	4.75	-2.75	5.26	1.47	4.51	12.65	4.33
-22.69	6.73	-9.70	6.13	1.19	6.38	8.44	5.88	16.29	5.15
-19.36	7.68	-2.11	8.41	6.65	8.07	16.02	7.82	20.54	6.31
-16.02	9.14	4.86	11.42	12.41	9.60	24.51	9.99	26.60	7.58
-11.77	10.76	10.33	14.44	16.96	11.18	30.27	12.00	33.27	9.20
-7.22	12.79	17.61	21.28	24.23	14.43	37.85	15.64	38.43	11.17
-3.58	15.06	21.26	27.47	28.18	18.06	46.34	22.81	43.89	13.69
0.36	18.10	24.61	38.08	32.43	22.84	51.21	32.27	51.16	18.21
4.31	21.52			36.69	29.47	51.82	36.48	54.50	21.44
7.35	26.94			37.91	36.89			59.36	28.24
10.09	32.37							62.11	37.97
12.21	37.72								

Source: Uppaluri 2010

Table C.18: Variation of Kf (Flooding factor) for various tray and sieve specifications.

Sieve & Valve tray flood line		Sieve and valve tray design		Bubble cap trays (flood)	
Tray spacing (inches)	K_f	Tray spacing (inches)	K_f	Tray spacing (inches)	K_f
10.1	309.1	13.4	505.7	9.9	151.5
10.6	372.4	14.2	580.7	11.3	226.8
11.2	441.5	15.0	635.5	12.4	304.7
12.0	507.8	16.1	713.5	13.2	359.6
13.2	605.9	17.2	774.3	14.7	480.7
14.7	712.7	18.7	835.3	15.6	538.5
16.2	828.1	20.1	902.0	17.2	596.7
17.4	909.0	22.1	980.4	18.4	643.3
18.9	964.3	23.3	1032.7	20.0	712.9
19.9	1010.7	25.0	1082.3	23.4	786.4
21.7	1063.3	26.4	1117.5	25.1	816.1
23.7	1116.0	28.5	1155.9	28.8	884.0
25.0	1156.8	30.1	1185.5	30.3	904.9
27.1	1212.4	32.2	1203.9	32.2	917.5
28.7	1256.3	35.0	1225.5	33.6	926.9
30.1	1294.4				
30.9	1317.7				

Source: Uppaluri 2010

Table C.19: Steam table data.

Psia (psi)	Saturation Temp. (°F)	Enthalpy (Btu/lb) at temperature										
		350 °F	400 °F	500 °F	600 °F	700 °F	800 °F	900 °F	1000 °F	1100 °F	1200 °F	1300 °F
15	213.03	1216.2	1239.9	1287.3	1335.2	1383.8	1433.2	1483.4	1534.5	1586.5	1693.2	1803.4
20	227.96	1215.4	1239.2	1286.9	1334.9	1383.5	1432.9	1483.2	1534.3	1586.3	1693.1	1803.3
30	250.34	1213.6	1237.8	1286.0	1334.2	1383.0	1432.5	1482.8	1534	1586.1	1692.9	1803.2
40	267.25	1211.7	1236.4	1285.0	1333.6	1382.5	1432.1	1482.5	1533.7	1585.8	1692.7	1803.0
50	281.02	1209.9	1234.9	1284.1	1332.9	1382.0	1431.7	1482.2	1533.4	1585.6	1692.5	1802.9
60	292.71	1208.0	1233.5	1283.2	1332.3	1381.5	1431.3	1481.8	1533.2	1585.3	1692.4	1802.8
70	302.93	1206.0	1232.0	1282.2	1331.6	1381.0	1430.9	1481.5	1532.9	1585.1	1692.2	1802.6
80	312.04	1204	1230.5	1281.3	1330.9	1380.5	1430.5	1481.1	1532.6	1584.9	1692	1802.5
90	320.28	1202	1228.9	1280.3	1330.2	1380	1430.1	1480.8	1532.3	1584.6	1691.8	1802.4
100	327.82	1199.9	1227.4	1279.3	1329.6	1379.5	1429.7	1480.4	1532	1584.4	1691.6	1802

Source: Uppaluri 2010

Table C.20: Characterization factor data table (Developed from correlation presented in Maxwell (1950)).

VABP	Characterization factor for various values of API																		
°F	90°	85°	80°	75°	70°	65°	60°	55°	50°	45°	40°	35°	30°	25°	20°	15°	10°	5°	0°
100	12.889	12.607	12.302	12.018	11.711	11.414	11.121	10.863	10.571	10.249	9.951	9.687	9.399						
120	13.040	12.755	12.451	12.157	11.859	11.564	11.255	10.983	10.695	10.382	10.080	9.802	9.504	9.202					
140	13.183	12.902	12.586	12.295	11.988	11.693	11.388	11.113	10.822	10.502	10.191	9.925	9.622	9.314					
160	13.331	13.042	12.725	12.426	12.125	11.831	11.519	11.236	10.938	10.627	10.319	10.042	9.728	9.419					
180	13.465	13.177	12.855	12.550	12.252	11.955	11.641	11.355	11.062	10.748	10.432	10.148	9.836	9.527	9.199				
200	13.600	13.314	12.995	12.686	12.391	12.086	11.759	11.474	11.181	10.859	10.541	10.257	9.937	9.631	9.317				
220	13.735	13.447	13.130	12.819	12.514	12.205	11.879	11.587	11.264	10.969	10.645	10.360	10.036	9.727	9.419				
240	13.863	13.577	13.255	12.941	12.624	12.330	11.998	11.692	11.396	11.079	10.757	10.466	10.133	9.826	9.515				
260	13.982	13.703	13.382	13.069	12.744	12.440	12.118	11.796	11.501	11.186	10.858	10.564	10.230	9.919	9.612	9.188			
280		13.828	13.504	13.187	12.854	12.555	12.221	11.904	11.601	11.286	10.948	10.655	10.325	10.003	9.703	9.301			
300		13.946	13.622	13.308	12.968	12.663	12.334	12.014	11.697	11.383	11.048	10.748	10.414	10.102	9.789	9.402			
320			13.742	13.428	13.082	12.780	12.444	12.117	11.792	11.481	11.138	10.840	10.497	10.193	9.879	9.486			
340			13.854	13.535	13.192	12.883	12.553	12.219	11.897	11.577	11.230	10.927	10.589	10.273	9.966	9.575	9.239		
360			13.959	13.650	13.298	12.977	12.655	12.320	11.992	11.671	11.326	11.018	10.677	10.353	10.047	9.666	9.324		
380				13.749	13.401	13.084	12.756	12.418	12.083	11.762	11.410	11.104	10.761	10.431	10.126	9.754	9.406		
400				13.854	13.514	13.181	12.862	12.514	12.183	11.855	11.498	11.184	10.843	10.524	10.206	9.833	9.489		
420				13.964	13.614	13.281	12.961	12.612	12.273	11.946	11.582	11.273	10.924	10.608	10.288	9.915	9.573	9.183	
440					13.724	13.386	13.056	12.712	12.364	12.030	11.665	11.358	11.002	10.682	10.365	9.996	9.649	9.283	
460					13.821	13.490	13.156	12.809	12.460	12.112	11.760	11.436	11.079	10.758	10.436	10.072	9.726	9.365	
480					13.929	13.585	13.259	12.899	12.547	12.202	11.842	11.519	11.158	10.835	10.506	10.145	9.798	9.444	
500						13.677	13.347	12.988	12.635	12.292	11.929	11.595	11.236	10.912	10.579	10.212	9.867	9.521	
520						13.769	13.452	13.081	12.725	12.375	12.013	11.670	11.313	10.988	10.655	10.280	9.934	9.591	9.231
540						13.861	13.533	13.173	12.806	12.460	12.087	11.752	11.392	11.063	10.724	10.355	9.995	9.665	9.303
560						13.956	13.628	13.261	12.903	12.543	12.162	11.830	11.470	11.140	10.788	10.425	10.068	9.736	9.371
580							13.717	13.348	12.985	12.627	12.256	11.902	11.541	11.218	10.863	10.489	10.137	9.810	9.426

Source: Maxwell, 1950

APPENDIX D

D.0 Preliminary Desalter Design

The most important factor in selecting desalter configuration is emulsion formation. Roa 1990 reported that all crude contain moisture and salts to varying degrees. Water is likely to occur in emulsion form when the crudes are naphthenic or sulferous. Onwuka 2010 also reported that one of the factors that may prevent two liquid phases from separating is due to droplets carrying electric charges due to dissolved ions, and these charges can cause the droplets to repel each other rather than coalesce into larger particles and settle by gravity. Langevin et al 2004 reported that the high asphaltene contain of heavy crude contribute to emulsion formation. So judgement as to:

Is chemical treatment required and to what extent?

Is heat treatment required and to what extent?

Is electrical treatment required and to what extent?

These are based on emulsion formation of the water - oil mix. It is worthy of note that the separation stage is the slowest stage in desalting so is rate determining for the process. I e mixing and heating are fast and pose no problem to attain.

Basically our problem is getting right process for separation of water –oil which is a function of emulsion formation. Emulsion formation which is a function of salt and asphaltene content. Sulphur and asphaltene content of our crude are low which lessen worries for complexity of process. Also emulsion and separation can be handled with gravity but with more retention time. High retention time gives larger separator size which may not be economical. Heating and electricity reduces the retention time and resulting size of separator. Correlation between electric voltage and retention time has not being accessed but that between temperature and retention time is given below.

Table 2: Typical Retention Times for Liquid/Liquid Separation

Type of Separation (Hydrocarbon/Water Separators)	Retention Time
Above 35^0 API Hydrocarbon	3 to 5 min
Below 35^0 API Hydrocarbon	
100^0F and above	5 to 10 min
80^0F	10 to 20 min
60^0F	20 to 30 min

Source: Onwuka 2010.

It is also worthy of note that there is a trade-off between capital cost and operational cost. i. e. no heating – higher cost in fabricating a larger separator or heating – higher cost in heating crude water mix during operation. Economic analysis is not done in this work.

Derived from Stokes' law, capacity of separator is given below:

Vertical Vessels:

$$Wcl = C\left(\frac{Sh-Sl}{\mu}\right)(0.785)D_v^2 \quad - \quad - \quad 1$$

Horizontal Vessels:

$$Wcl = C\left(\frac{Sh-Sl}{\mu}\right)L_l H_l \quad - \quad - \quad - \quad 2$$

Separator volume is given as:

$$V = \frac{W(t)}{1440} - \quad - \quad - \quad - \quad - \quad 3$$

Wcl = flow rate of hydrocarbon Onwuka 2010

Sh = specific gravity of water

Sl = specific gravity of hydrocarbon

μ = viscosity of hydrocarbon

Dv = diameter of separator

L_l = length of separator

H_l = height of separator

V = volume of separator

W = total flow rate

t = retention time

C = constant given below.

Table 2: Values of C

Emulsion Characteristic	Droplet Diameter, Microns	Constant C
Free liquids	200	1100
Loose emulsion	150	619
Moderate emulsion	100	275
Tight emulsion	60	99

Source: Onwuka 2010.

Sizing of desalter.

Objective: to determine the size of a vertical desalter to handle 1 bpd of escravos crude with 20% water injected into crude.

Operating temperature = $60\text{-}80^0$F (room temperature)

Operating pressure = 15 psia (about atmospheric)

Water specific gravity = 1.0

Crude oil specific gravity = 0.86 (from Assay)

Crude oil viscosity =60.4 cp

Retention time = 30 min(from Table 2)

Based on information provided by literature earlier, crude emulsion characteristic is classified as free water. Therefore C = 1100

By Equation 1

$1 = 1100\frac{1-0.86}{60.4}(0.785)(D_v)^2$

D_v = 0.7ft = 0.21m

By Equation 3

$V = \frac{1.2 \times 30}{1440}$ = 0.017bbl = 3.975 dm^3

= 0.00398 m3

$V = \frac{H\pi D^2}{4}$ (volume of cylinder)

$0.00398 = \frac{H\pi 0.21^2}{4}$

H = 0.11 m

Size of desalter is 21cm OD x 15cm.

Design of feed storage tank

Objective: size rectangular tanks to hold water and crude oil feed for one hour run.

Crude feed per hour = 159/24 = 6.6 lit = 6600 cm^3

Water feed per hour = 0.2 x 6.6 = 1.3 lit = 1300 cm^3

Adding 300cm^3 and 200cm^3 to crude and water tank respectively to avoid flooding and some flexibility in operation.

Crude tank capacity = 6900 cm^3

Water tank capacity = 1500 cm^3

A workable dimension (H x B x L) is as:

Crude tank dimension = 30 x 10 x 23 cm

Water tank dimension = 30 x 10 x 5 cm

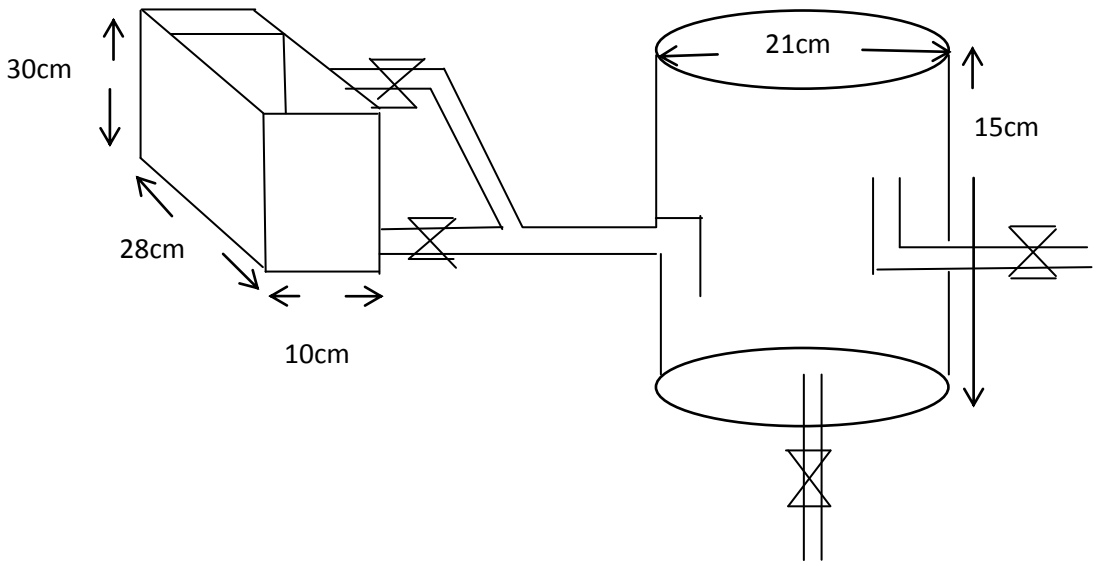


Figure E. 1: Working Drawing for Crude oil Desalter

APPENDIX E

E.0 Aspen Plus Run for Atmospheric Distillation Unit

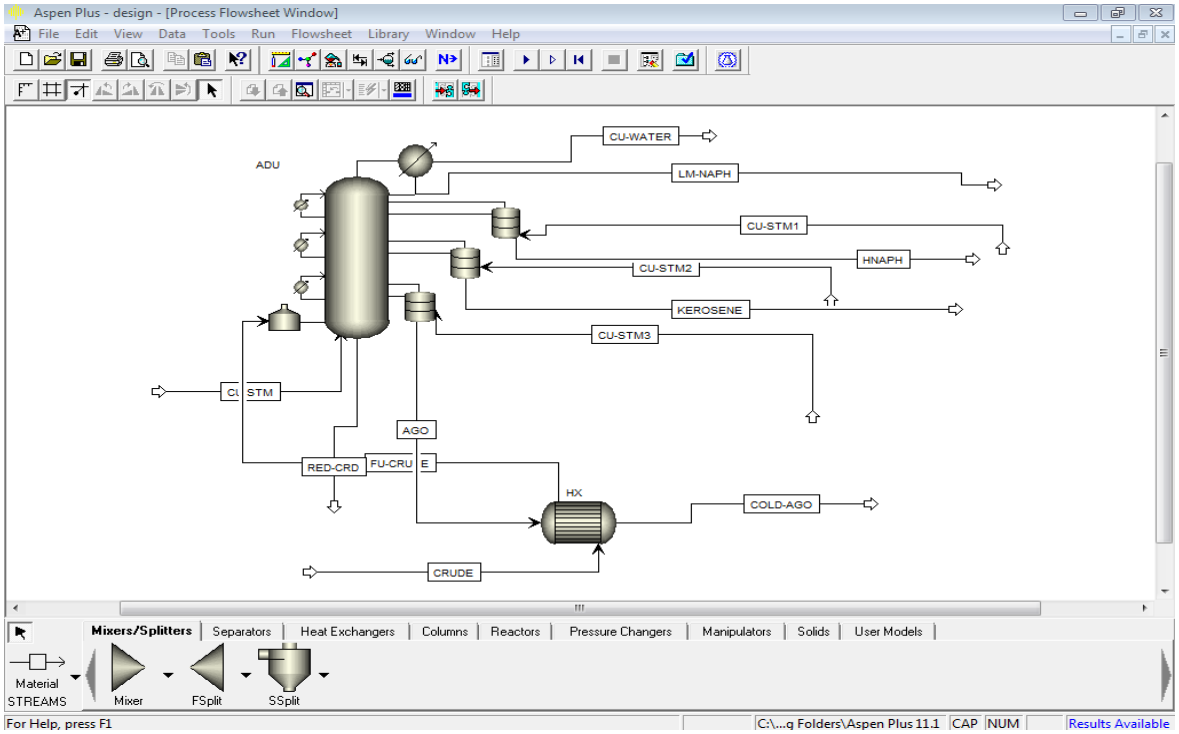


Figure E.1: Flowsheet for ADU

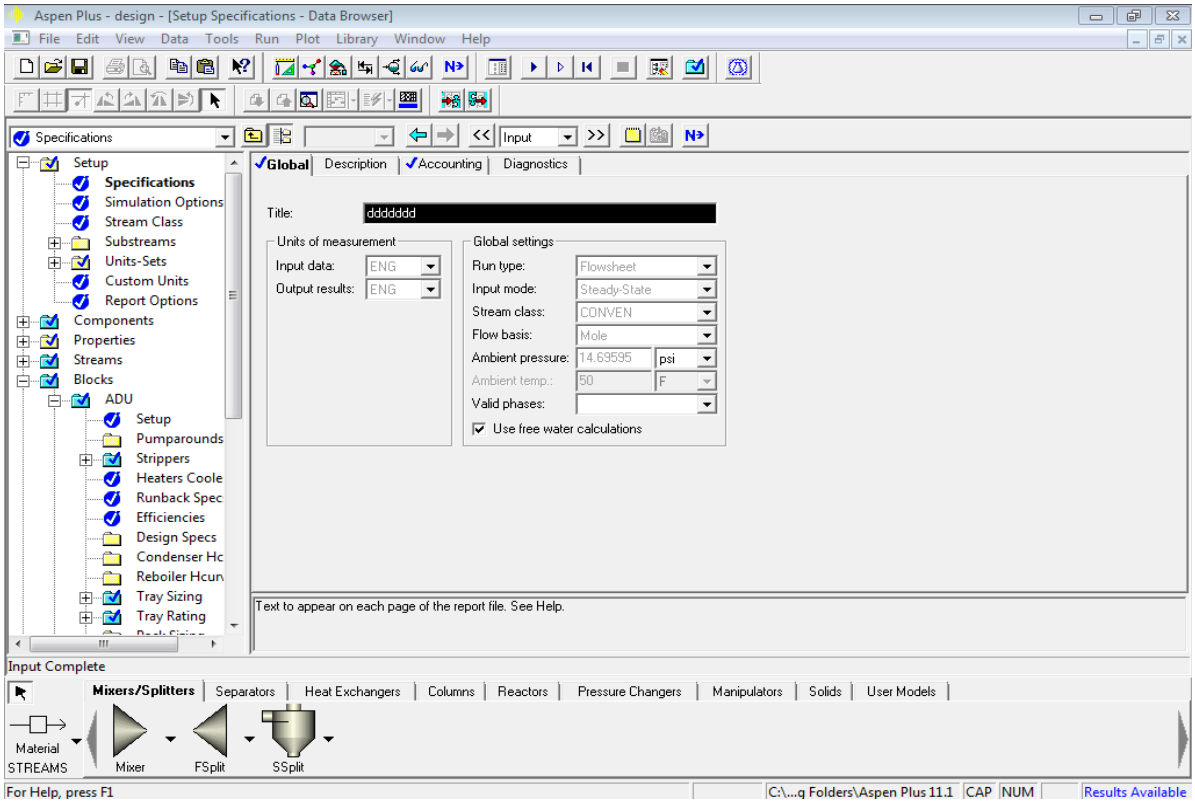


Figure E.2: Setup specification

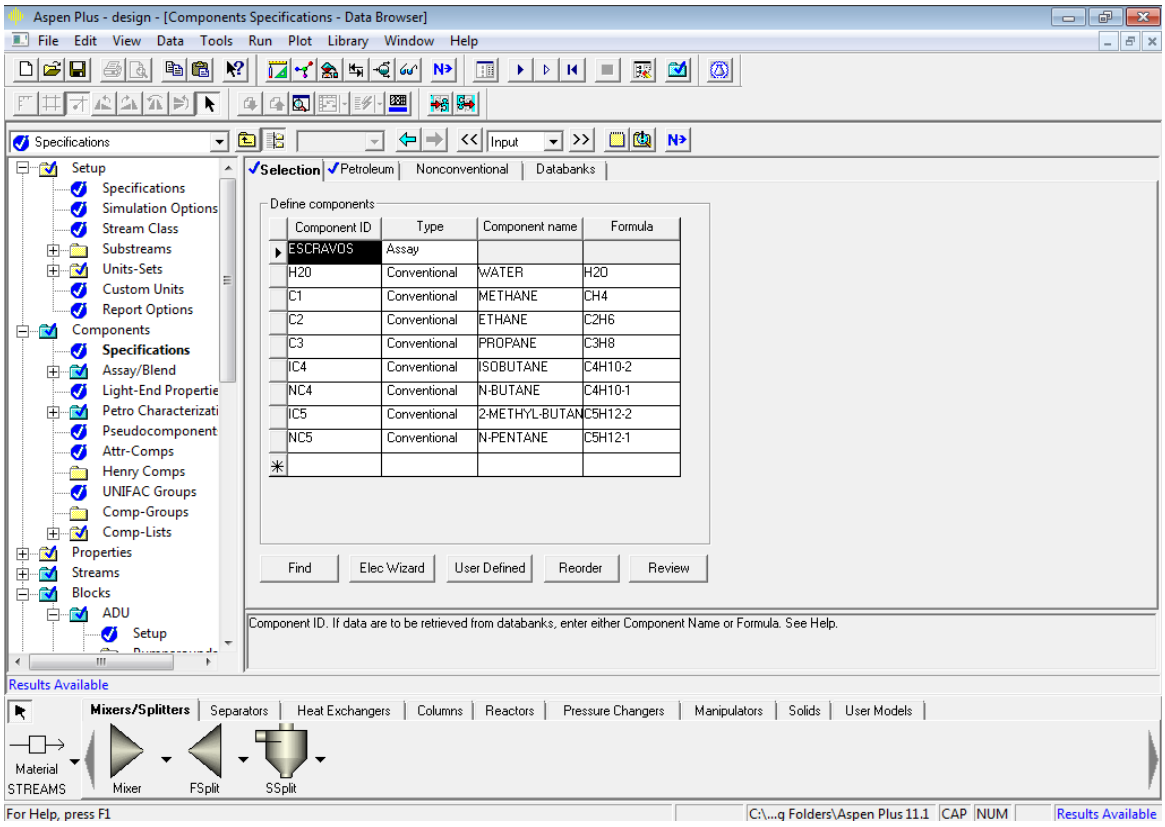


Figure E.3: Component Specification

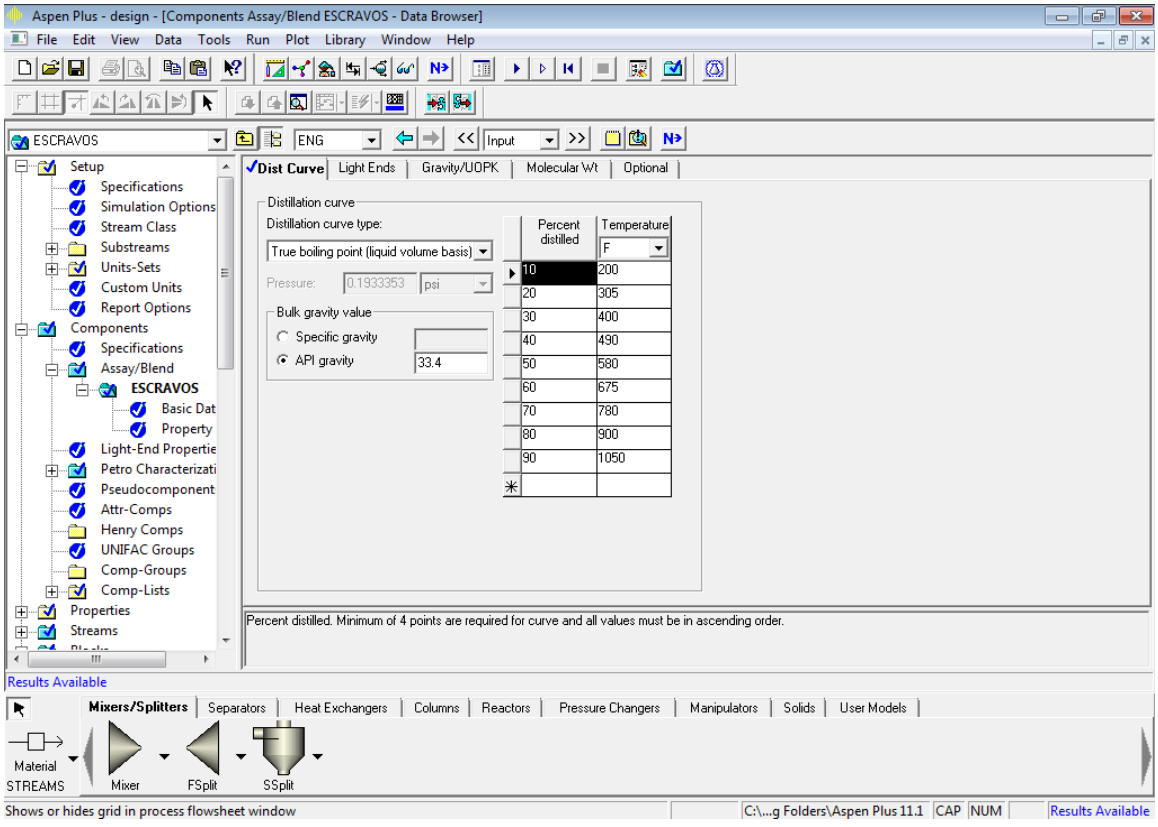


Figure E.4: Escravos crude Assay specification

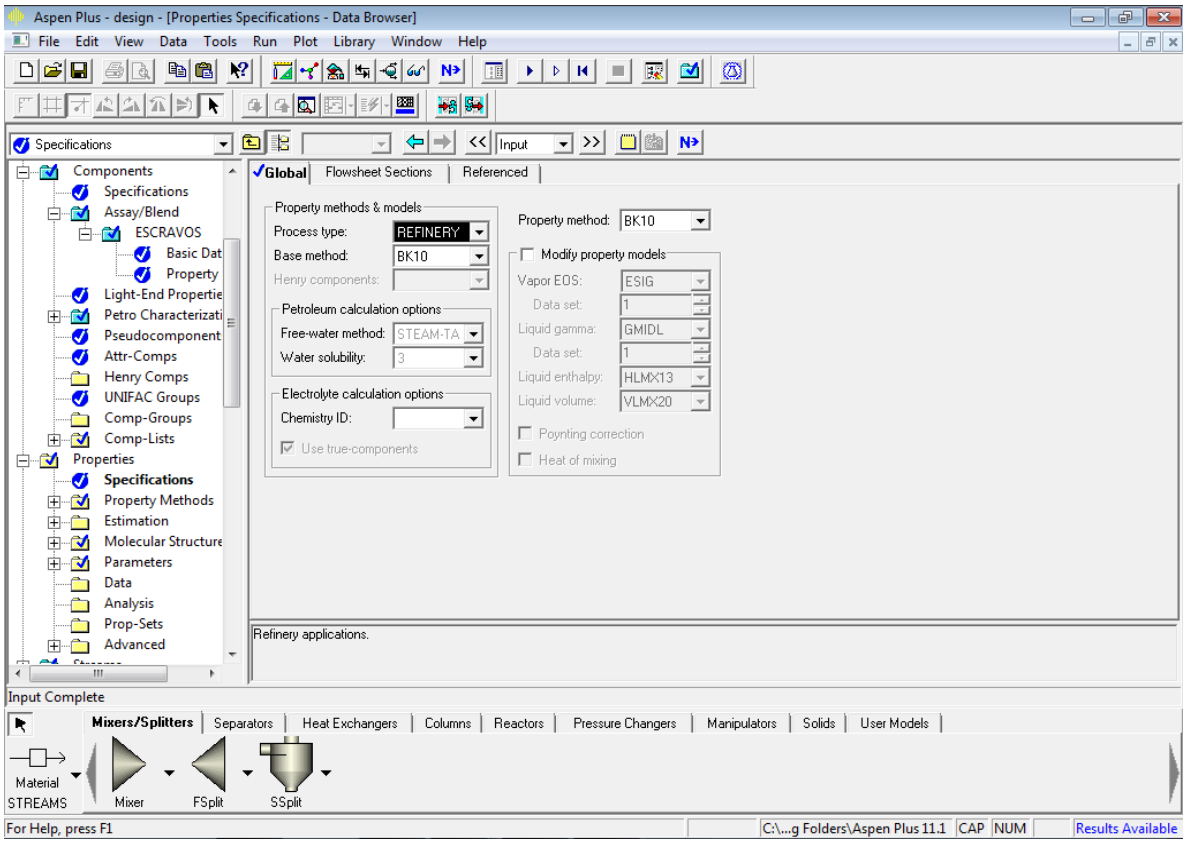


Figure E.5: Property specification

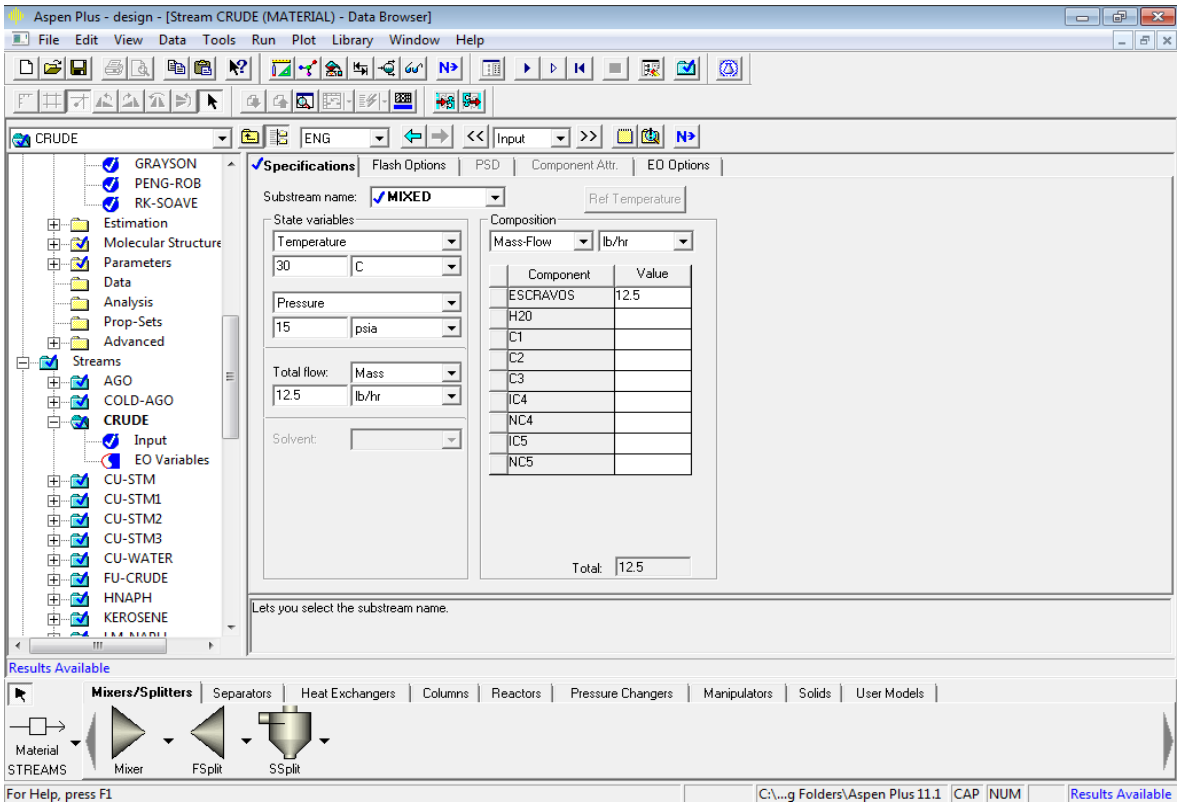


Figure E.6: Crude stream specification

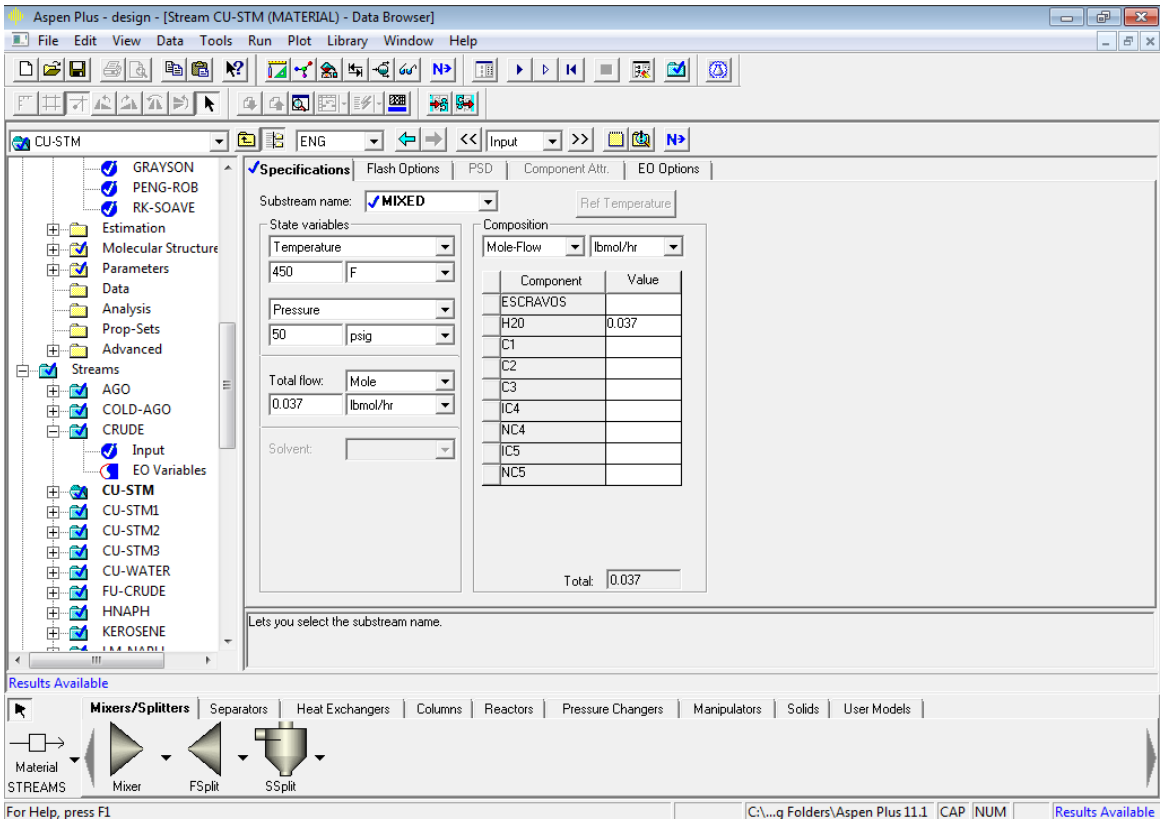


Figure E.7: specification of steam in main column

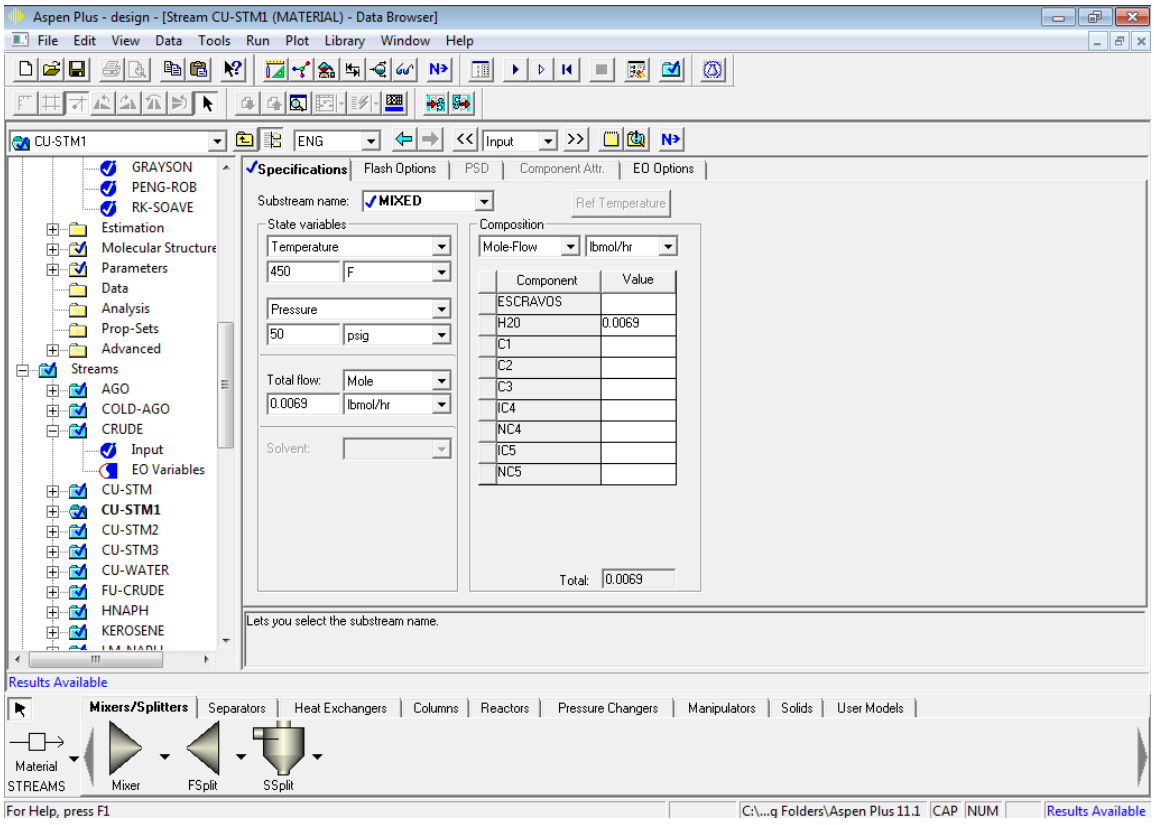


Figure E.8: Specification of steam to heavy naphtha stripper

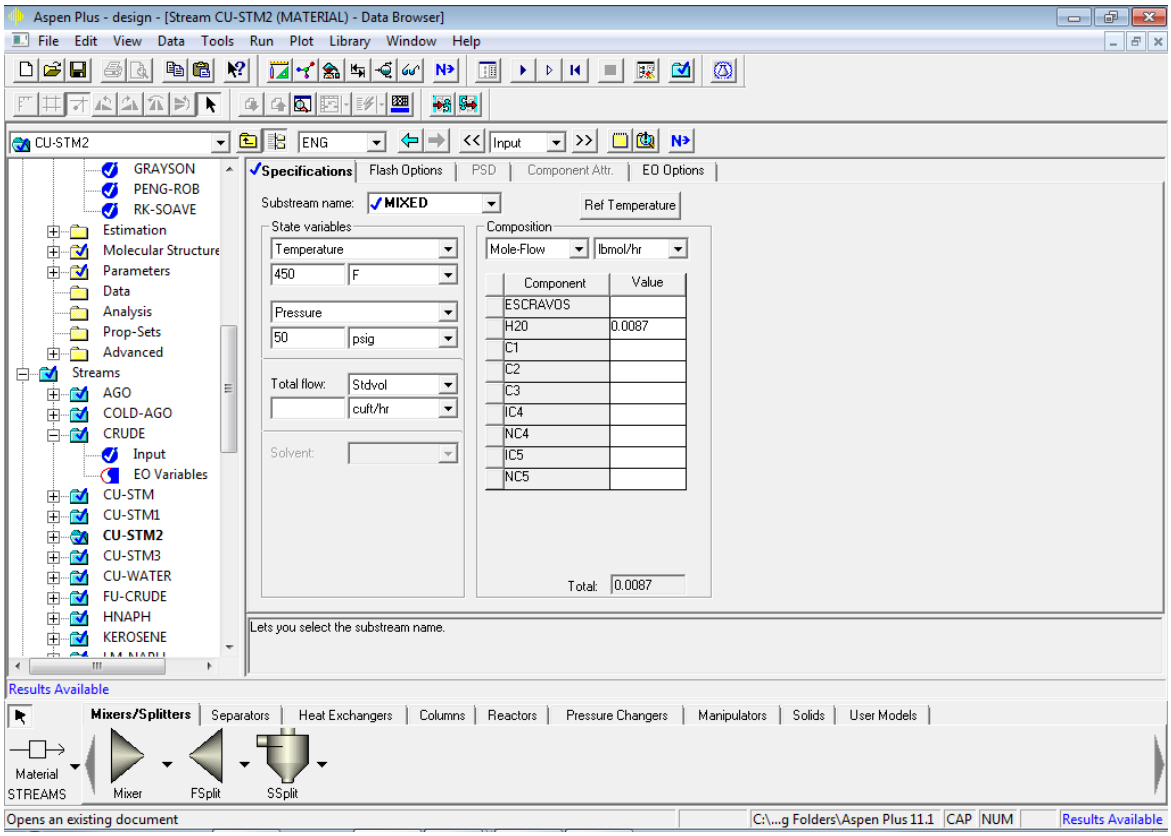


Figure E.9: Specification of steam to kerosene stripper

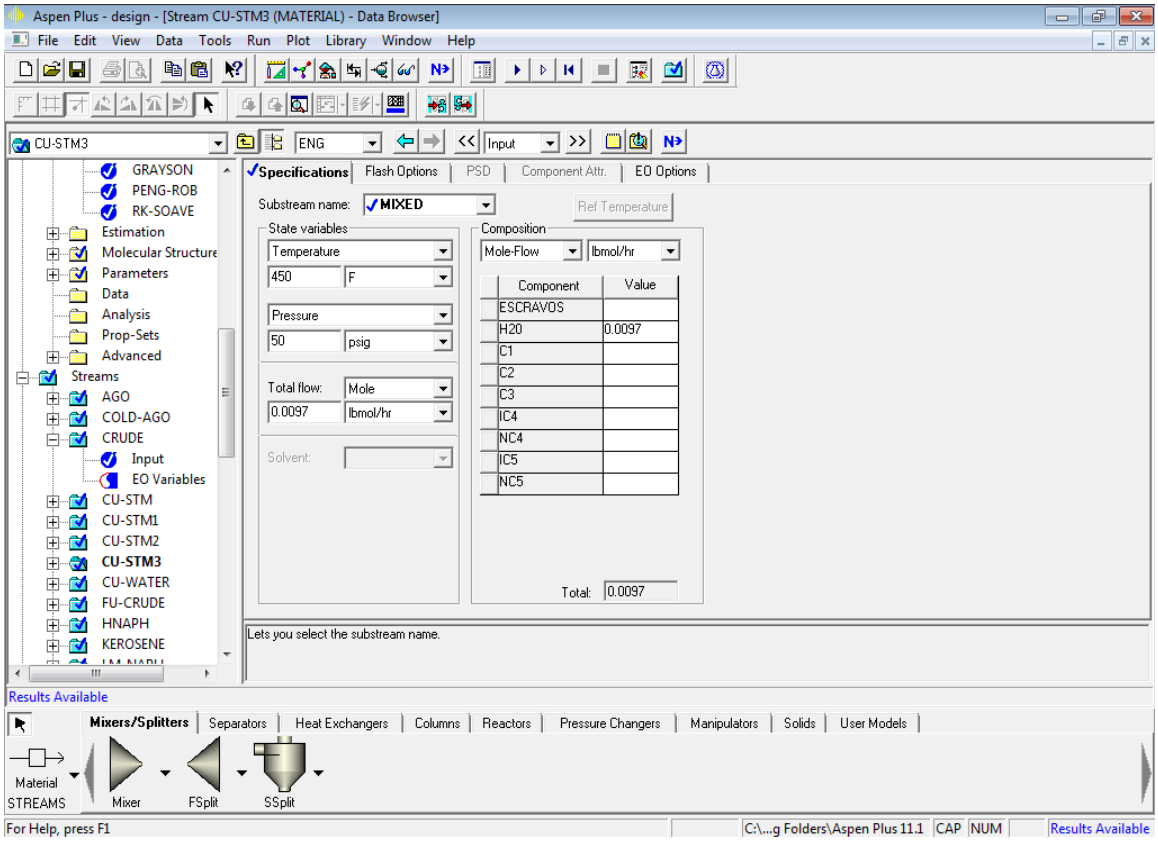


Figure E.10: Specification of steam to AGO stripper

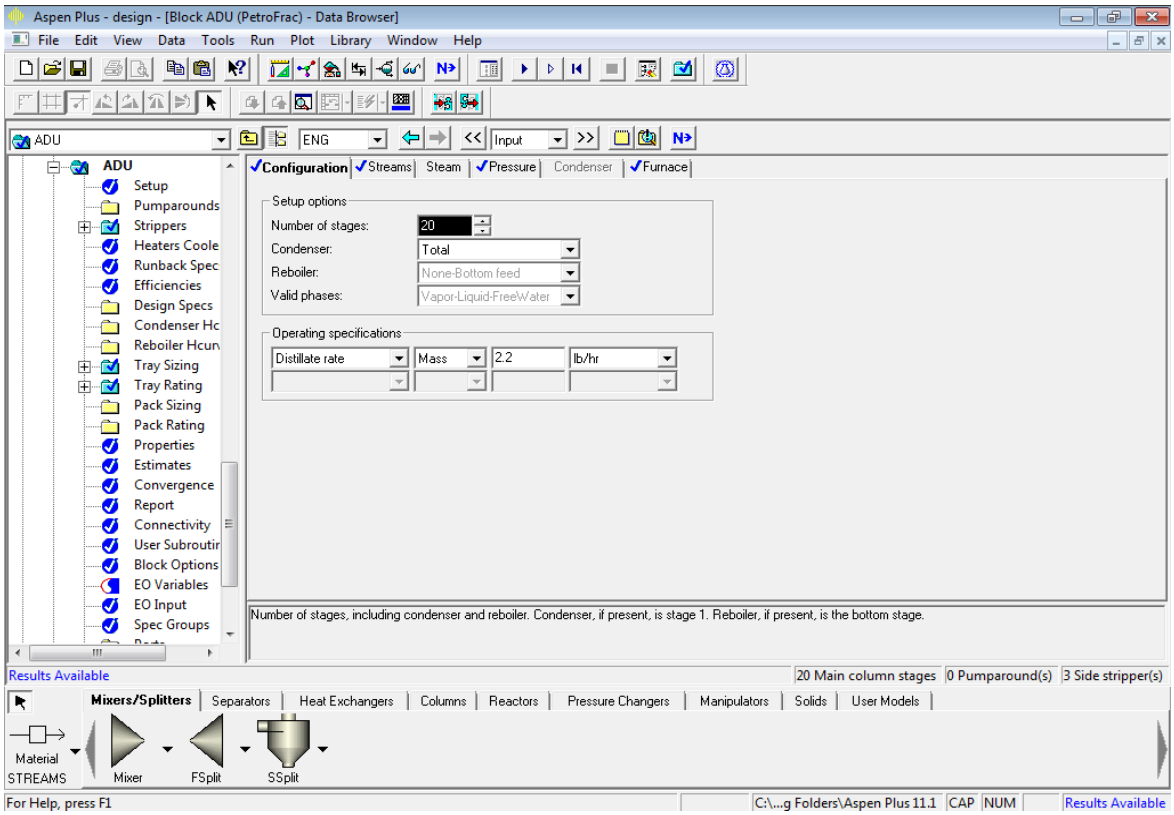


Figure E.11: Main column configuration

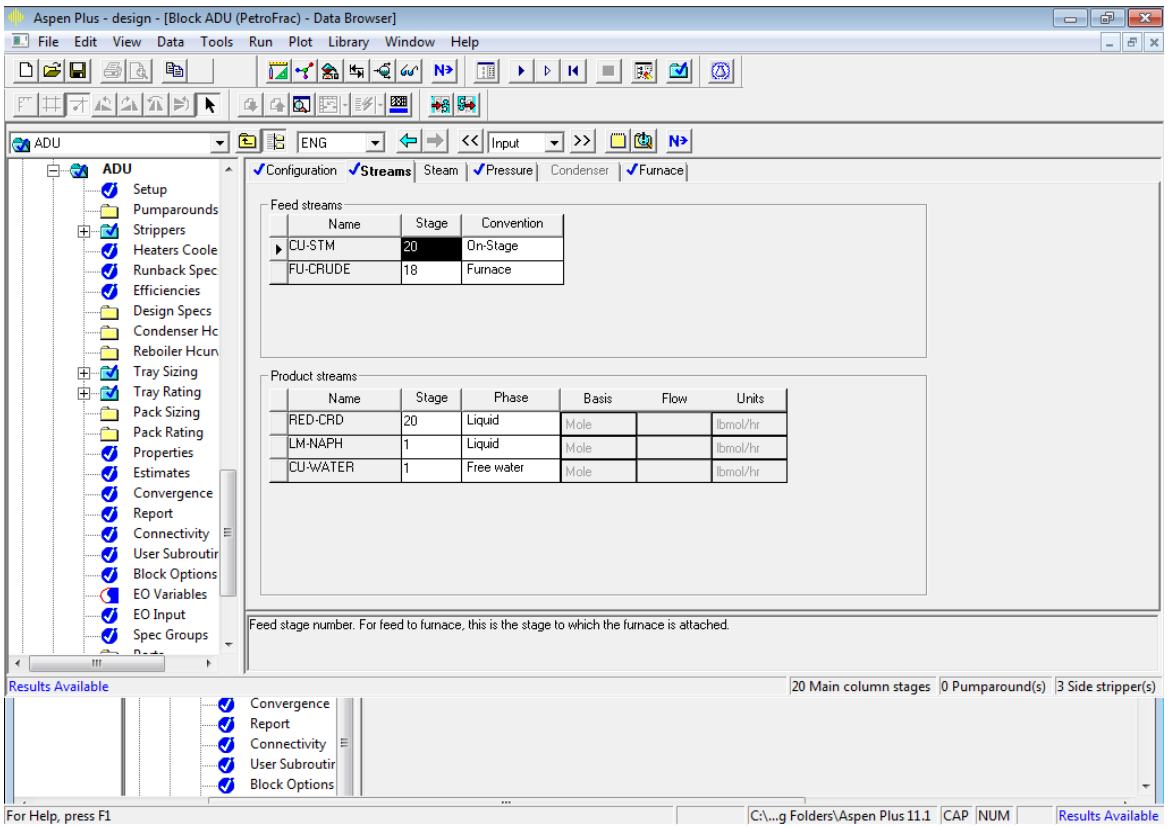


Figure E.12: Main column streams configuration

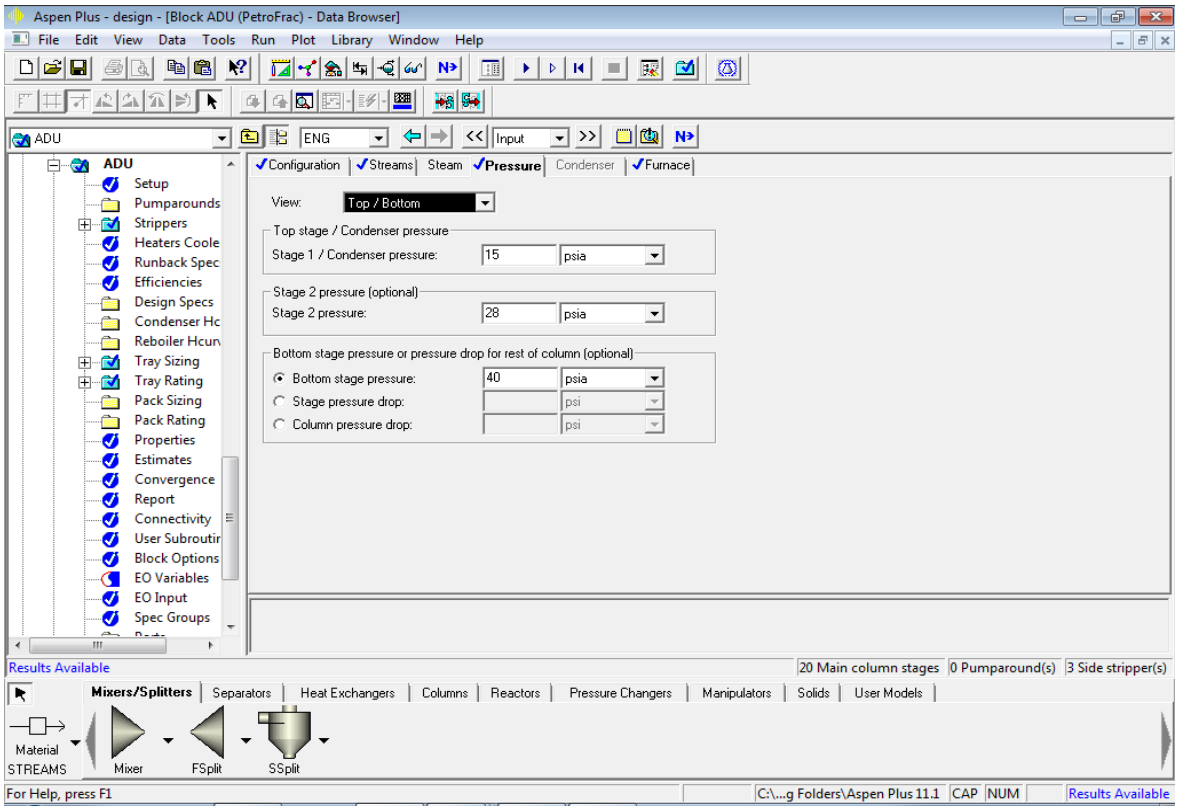


Figure E.13: main column pressure specifications

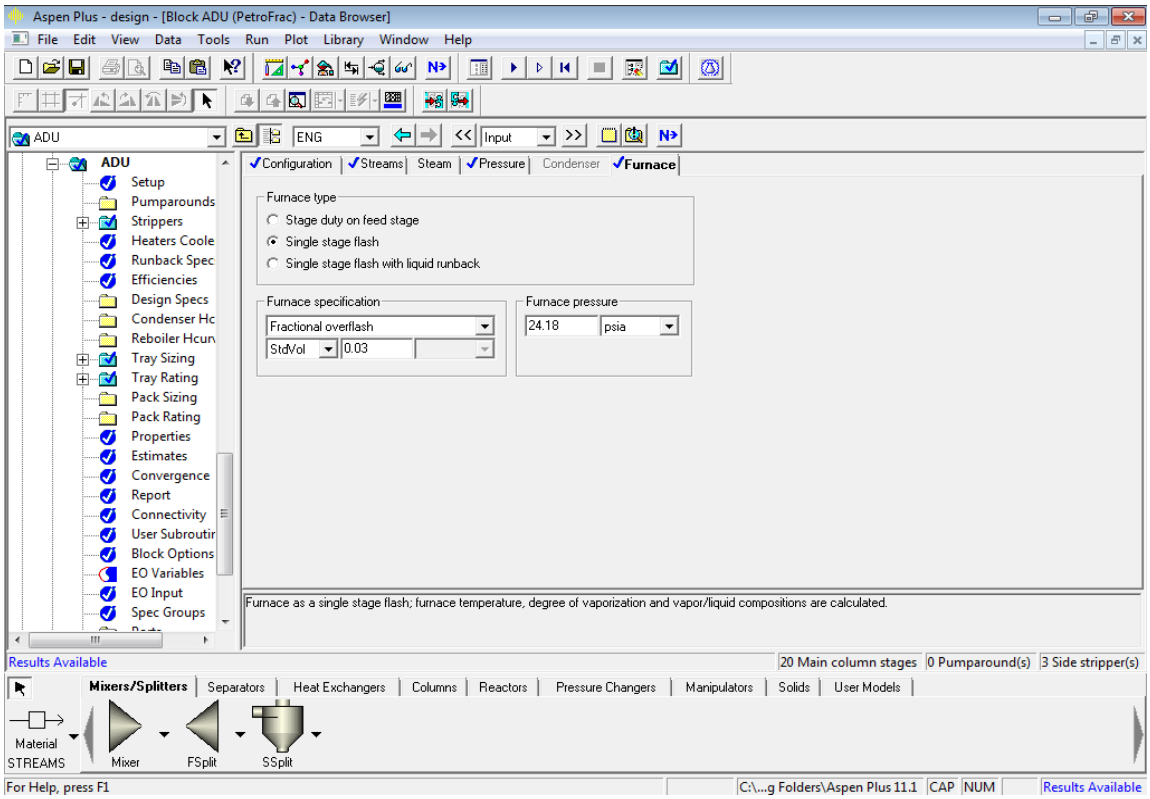


Figure E.14: Furnace specification

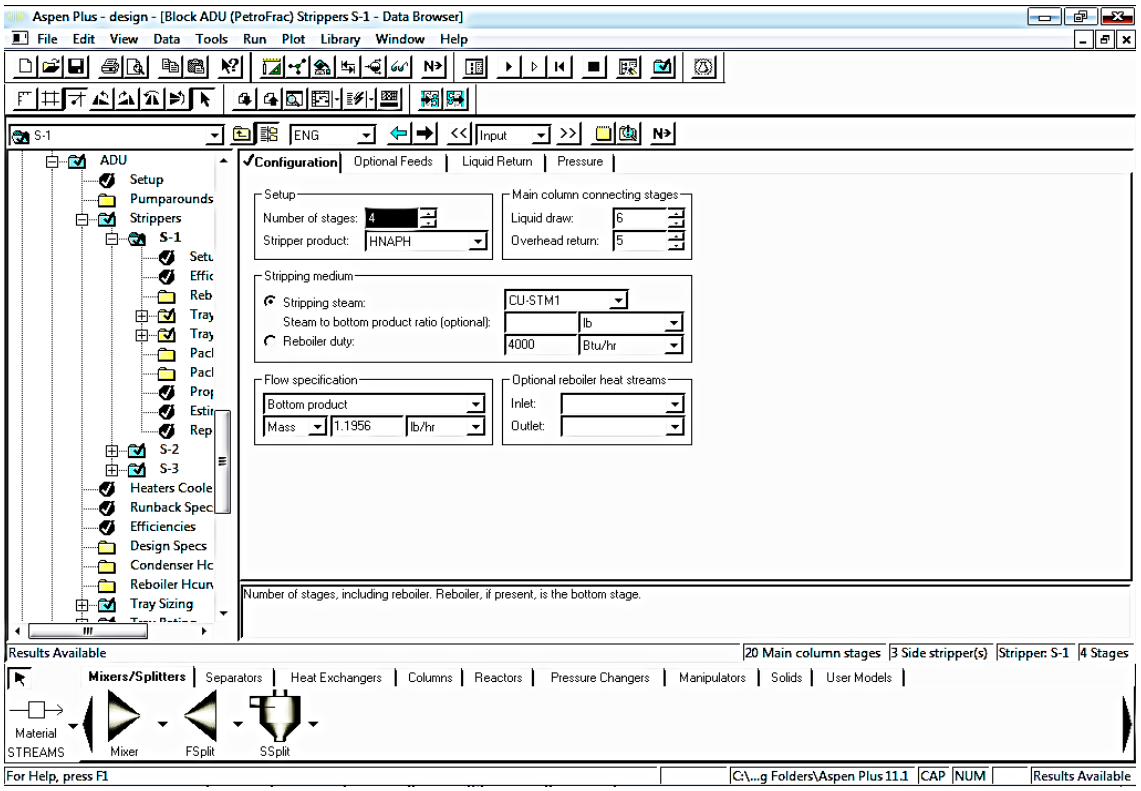


Figure E.15: heavy naphtha stripper specification

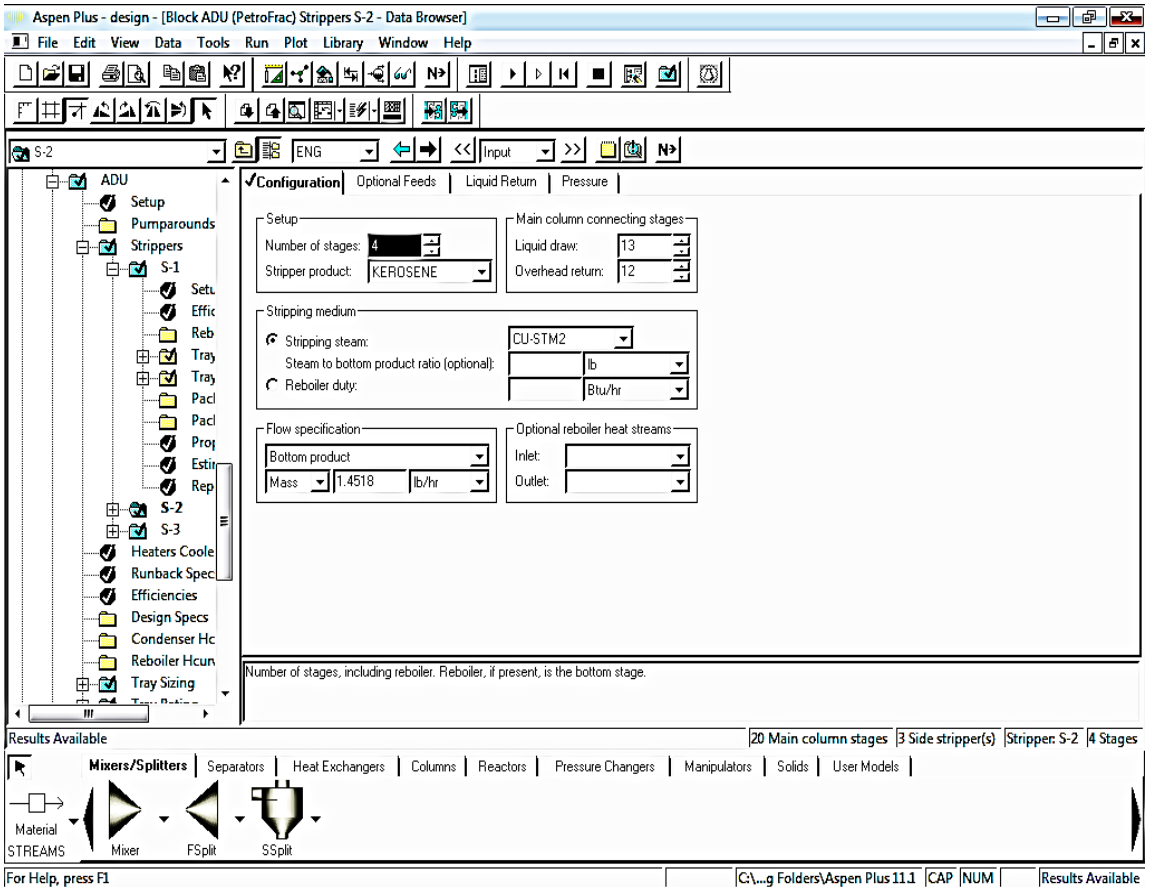


Figure E.16: Kerosene stripper specification

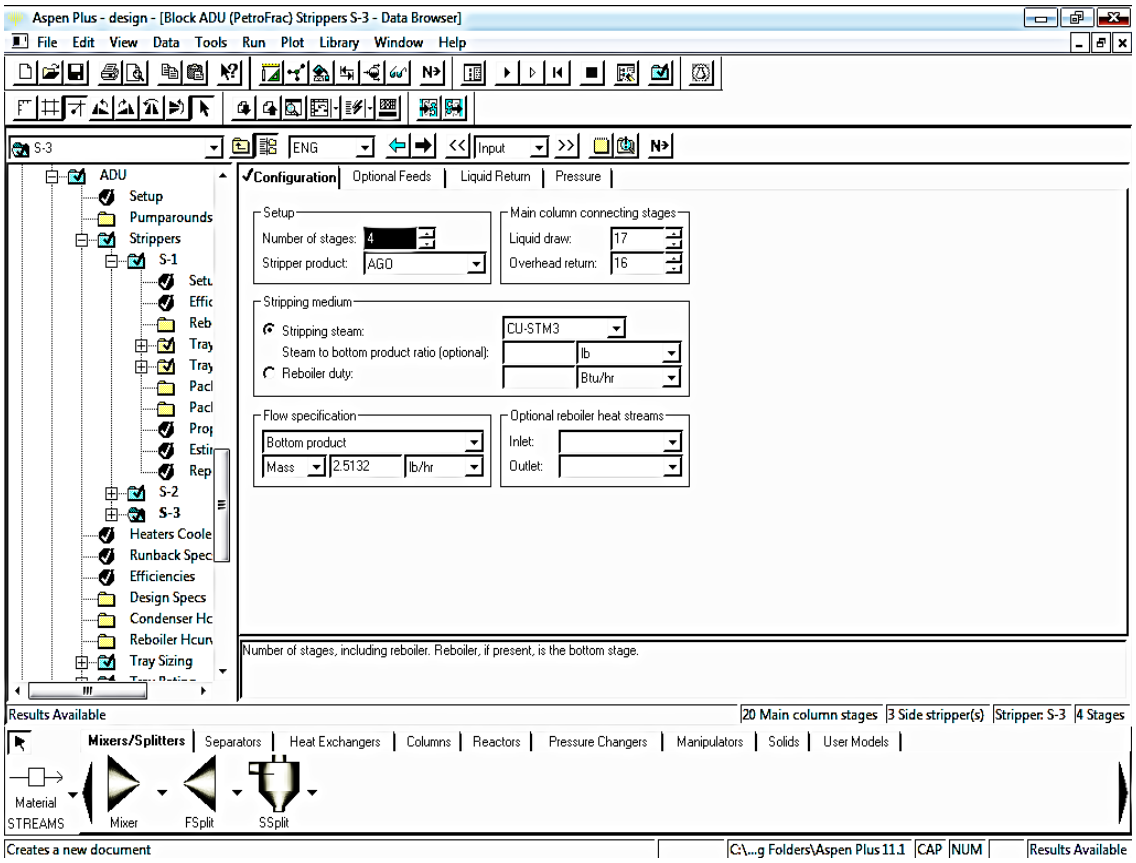


Figure E.17: AGO stripper specification

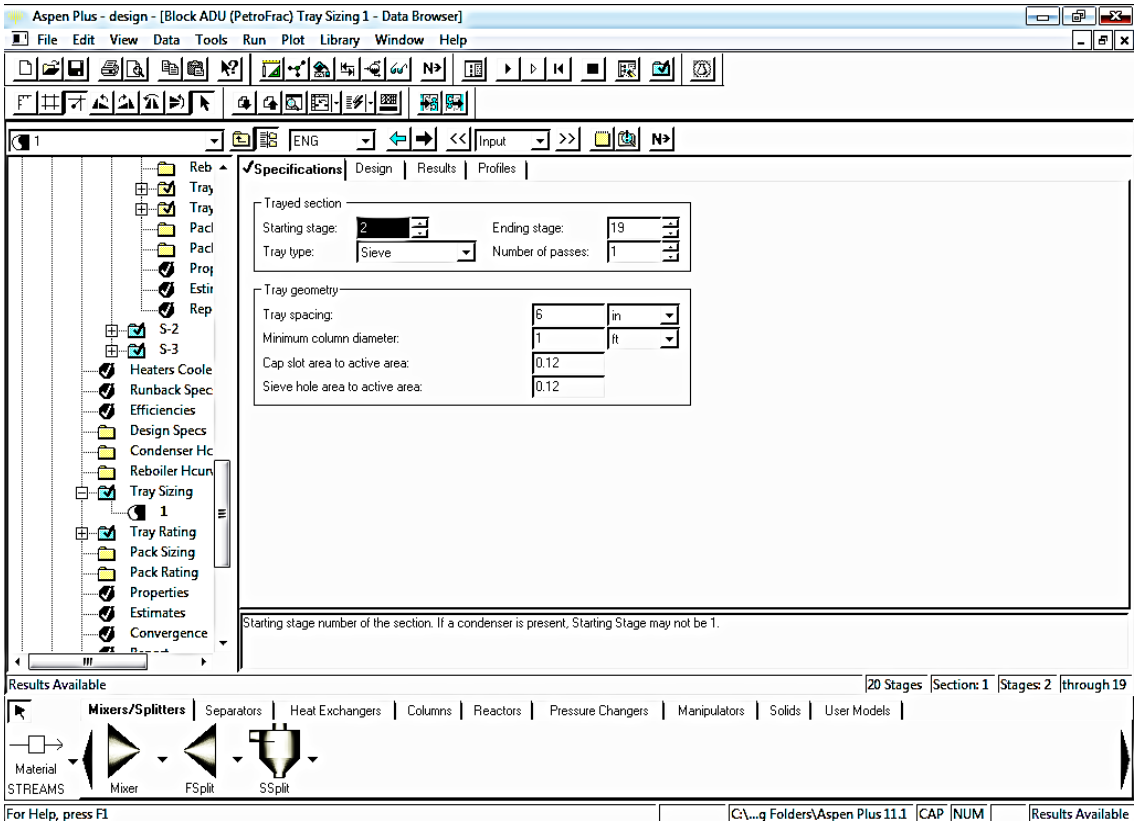


Figure E.18: ADU Tray sizing specification

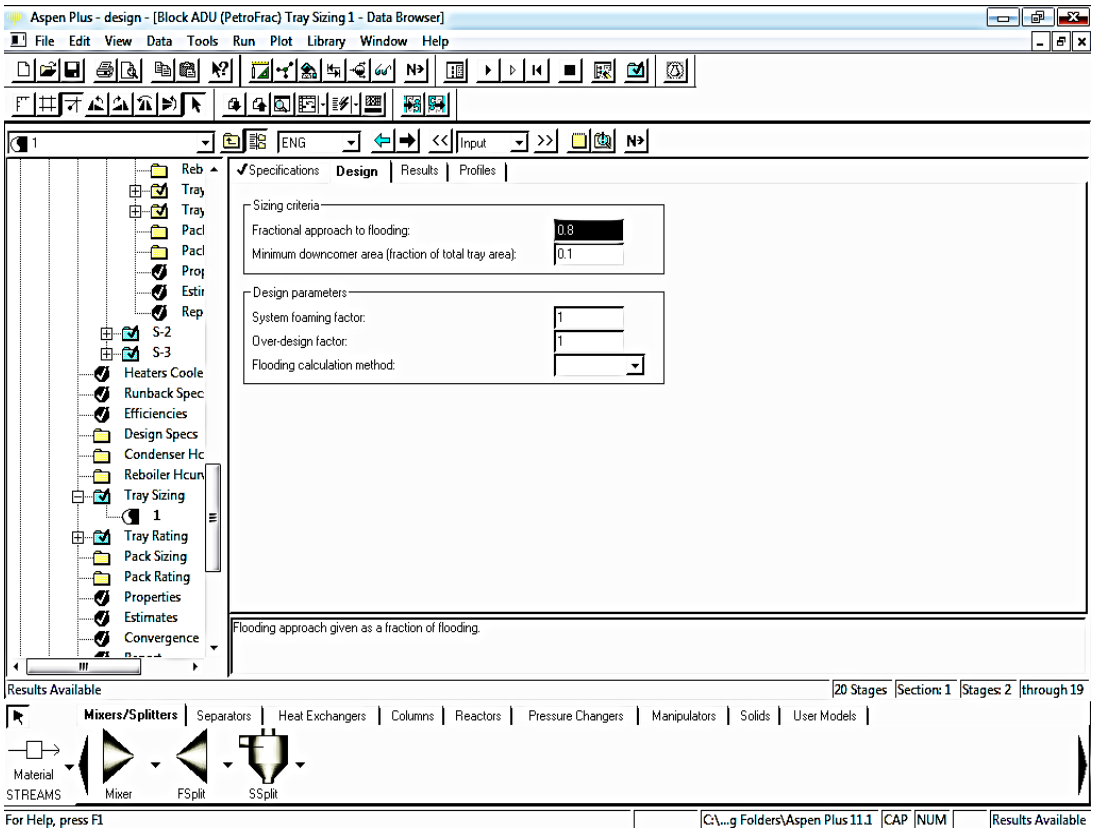


Figure E.19: ADU Tray sizing Design spec.

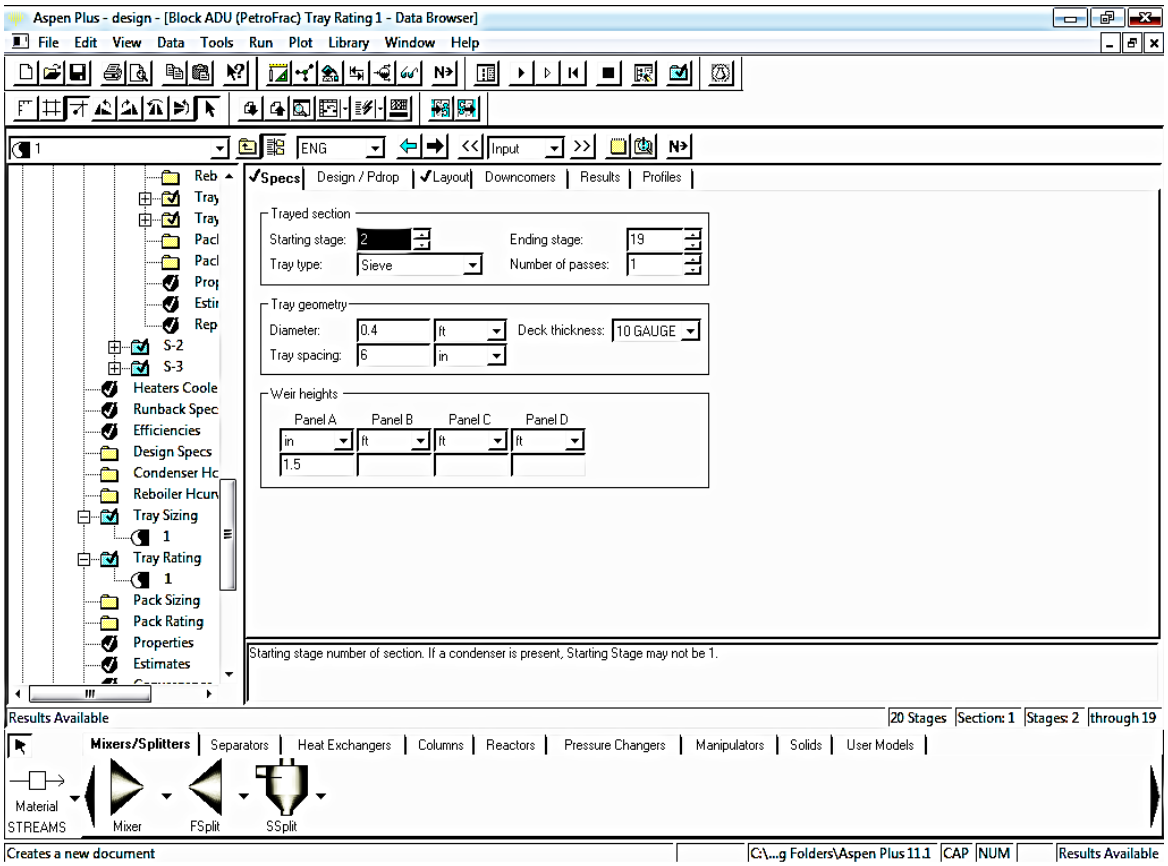


Figure E.20: ADU Tray rating specification

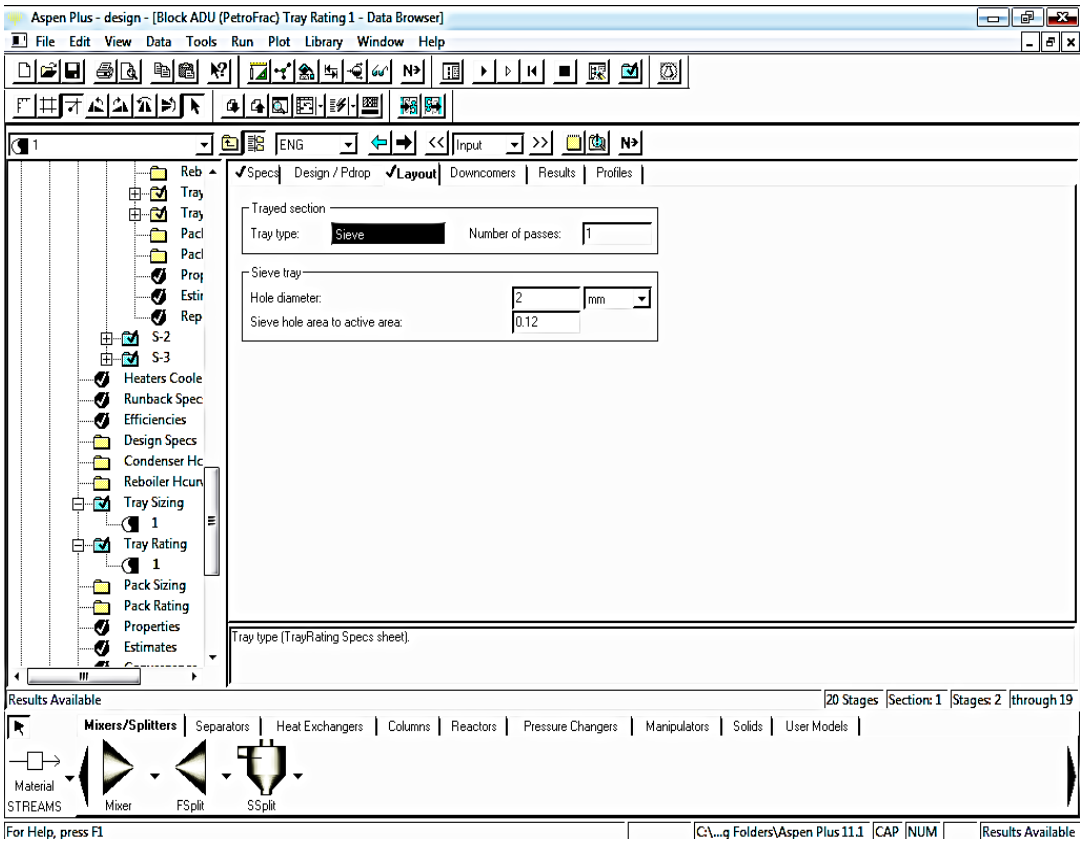


Figure E.21: ADU Tray rating layout

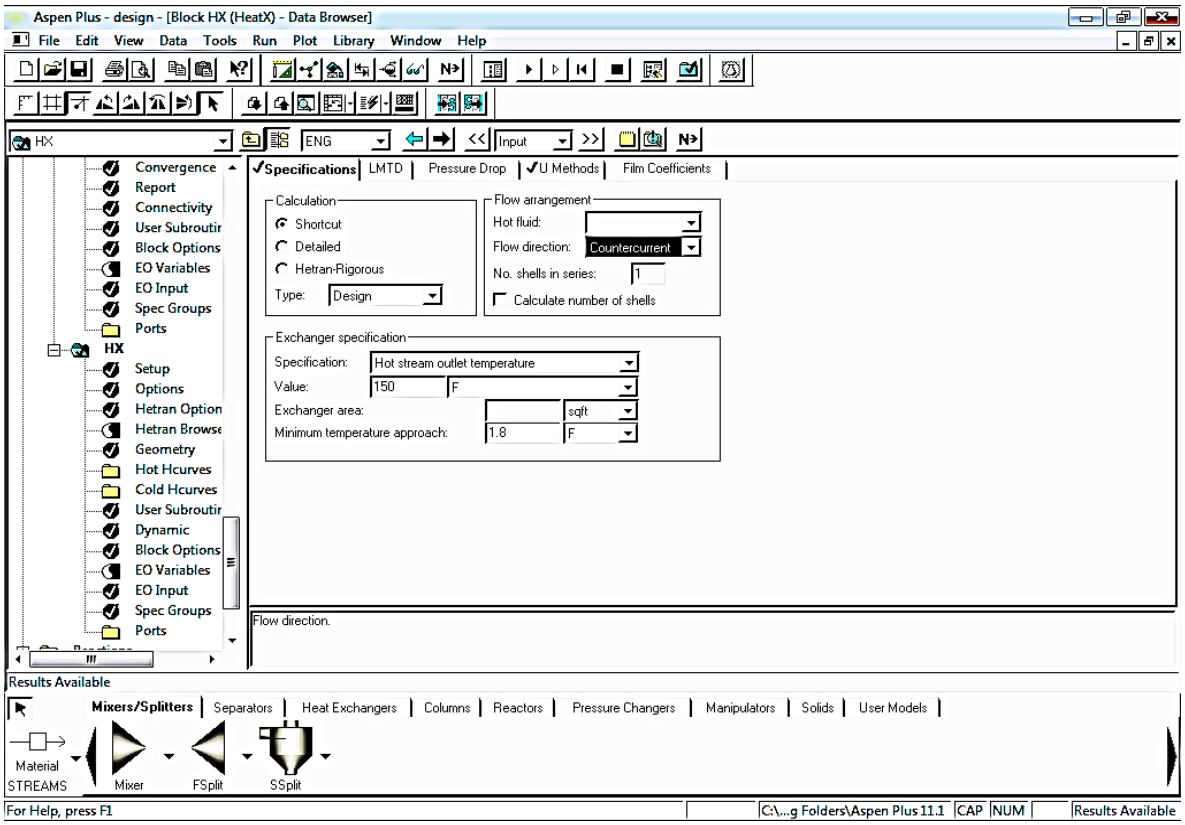


Figure E.22: Heat exchanger specification

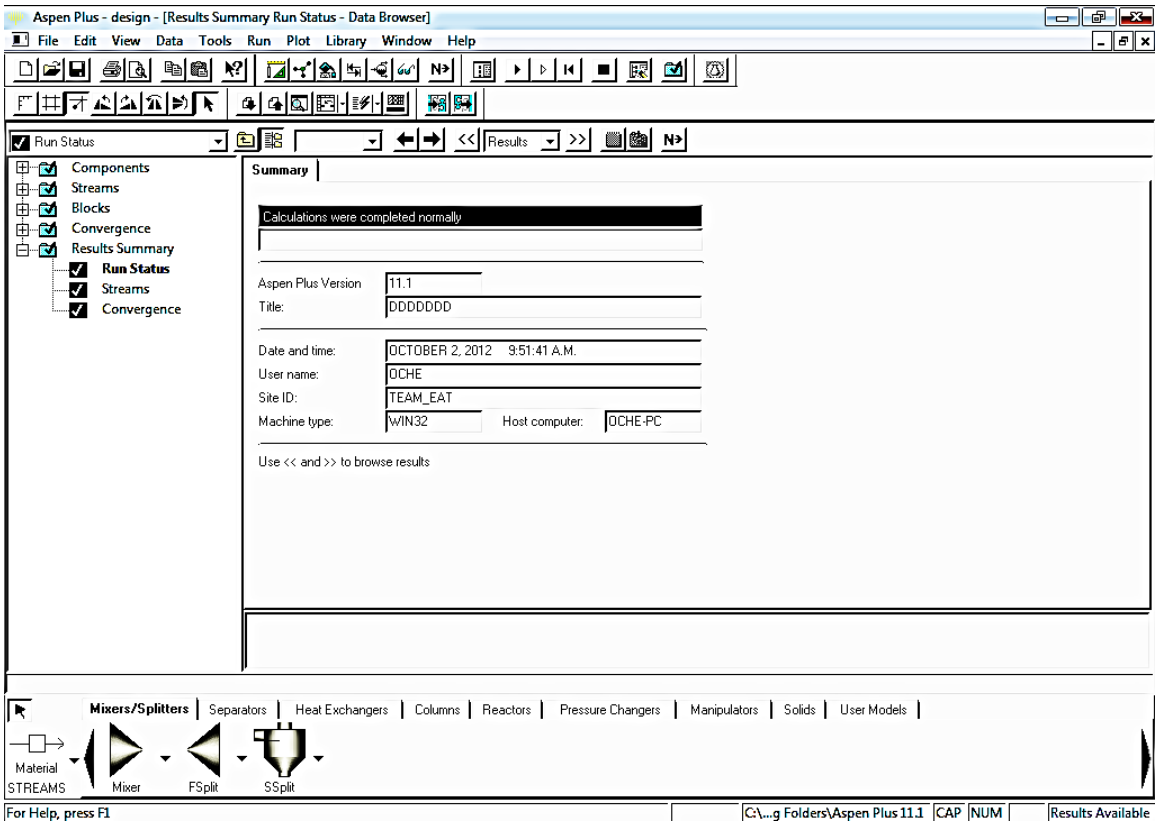


Figure E.23: ADU Result run status

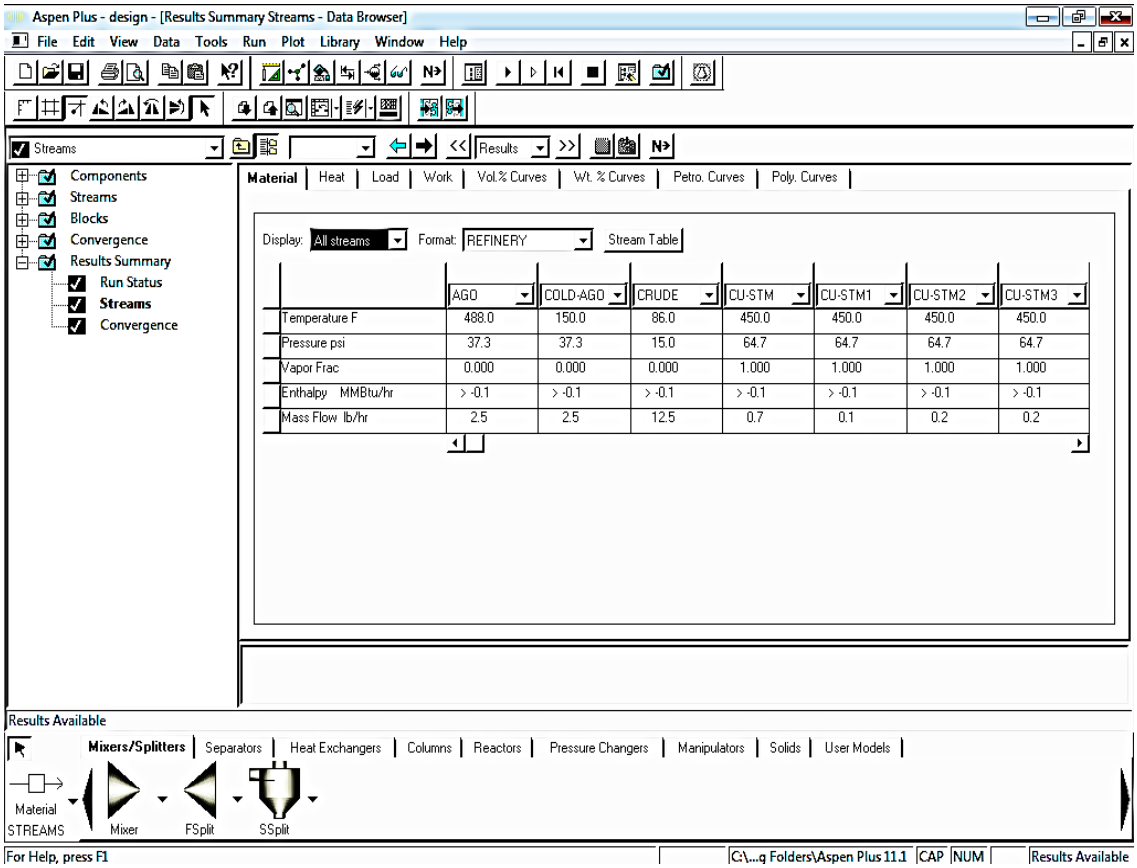


Figure E.24: ADU Stream results

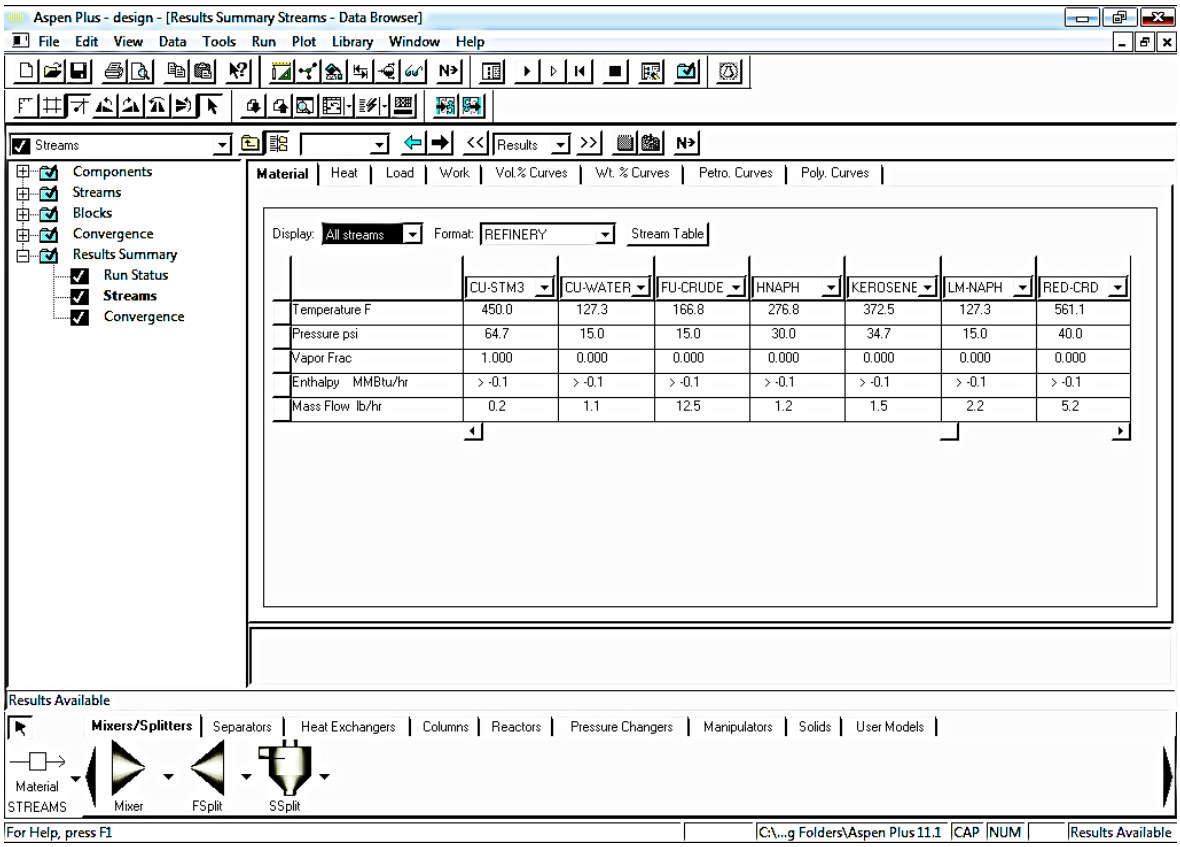


Figure E.25: ADU Stream results continues

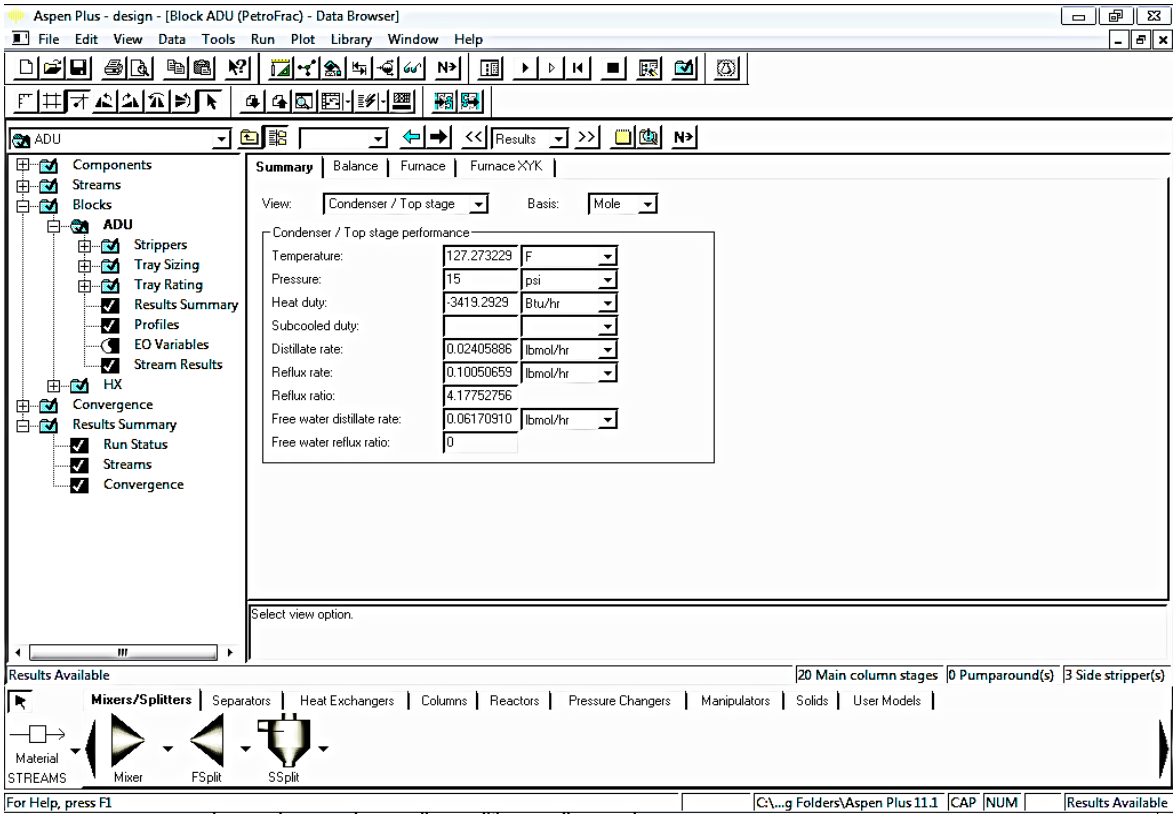


Figure E.26: Result summary of ADU

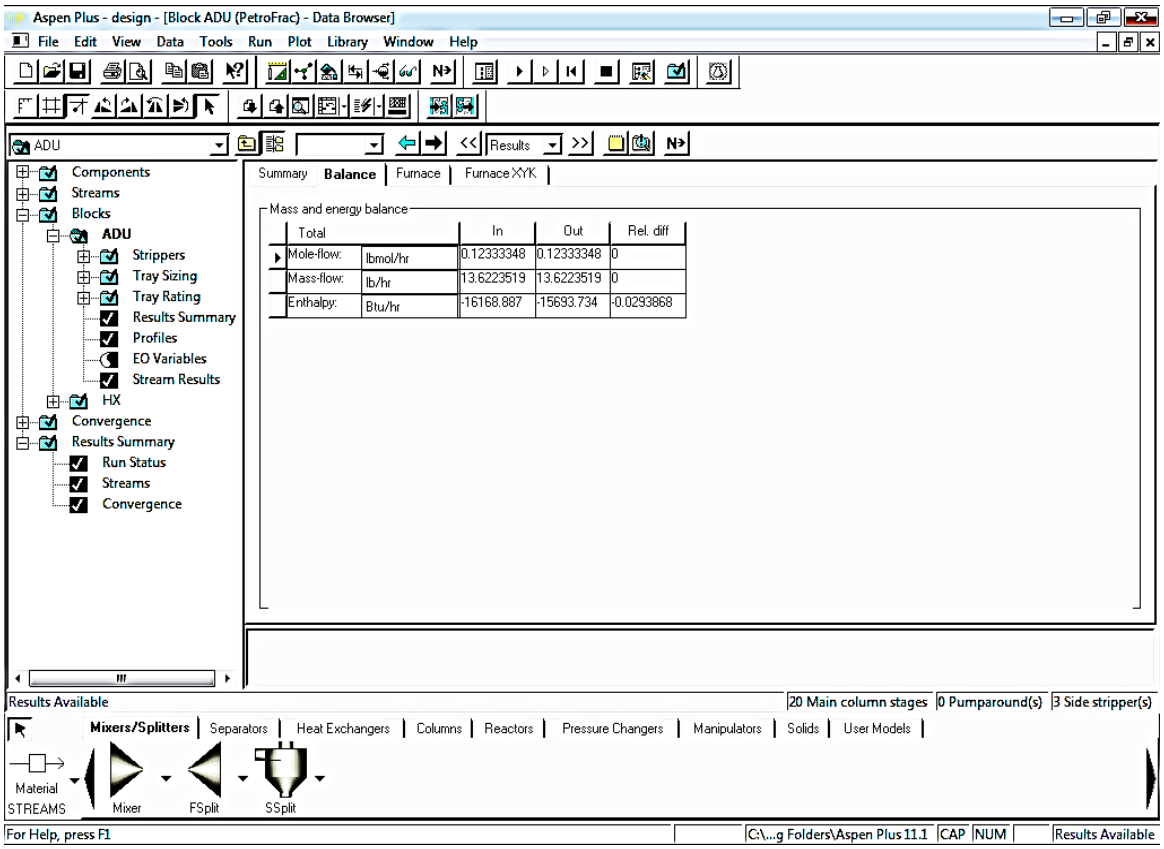


Figure E.27: ADU Material and energy balance res

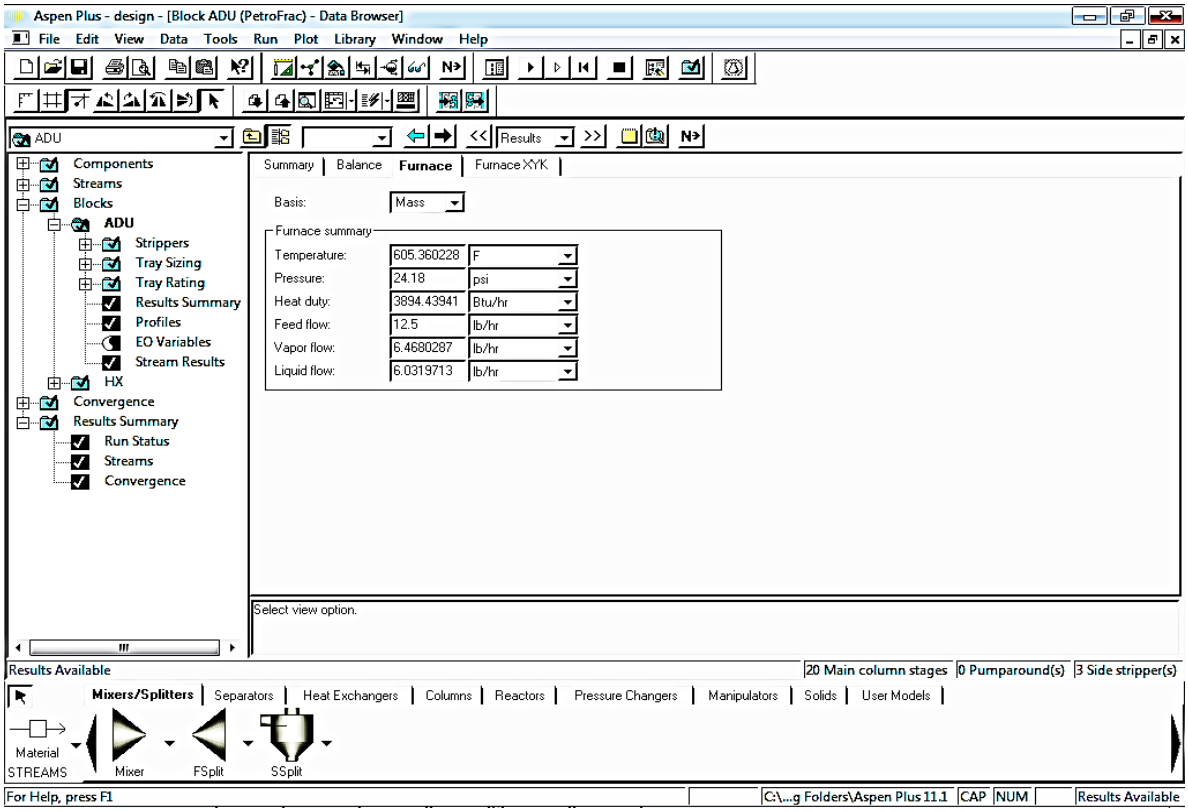


Figure E.28: ADU Furnace result

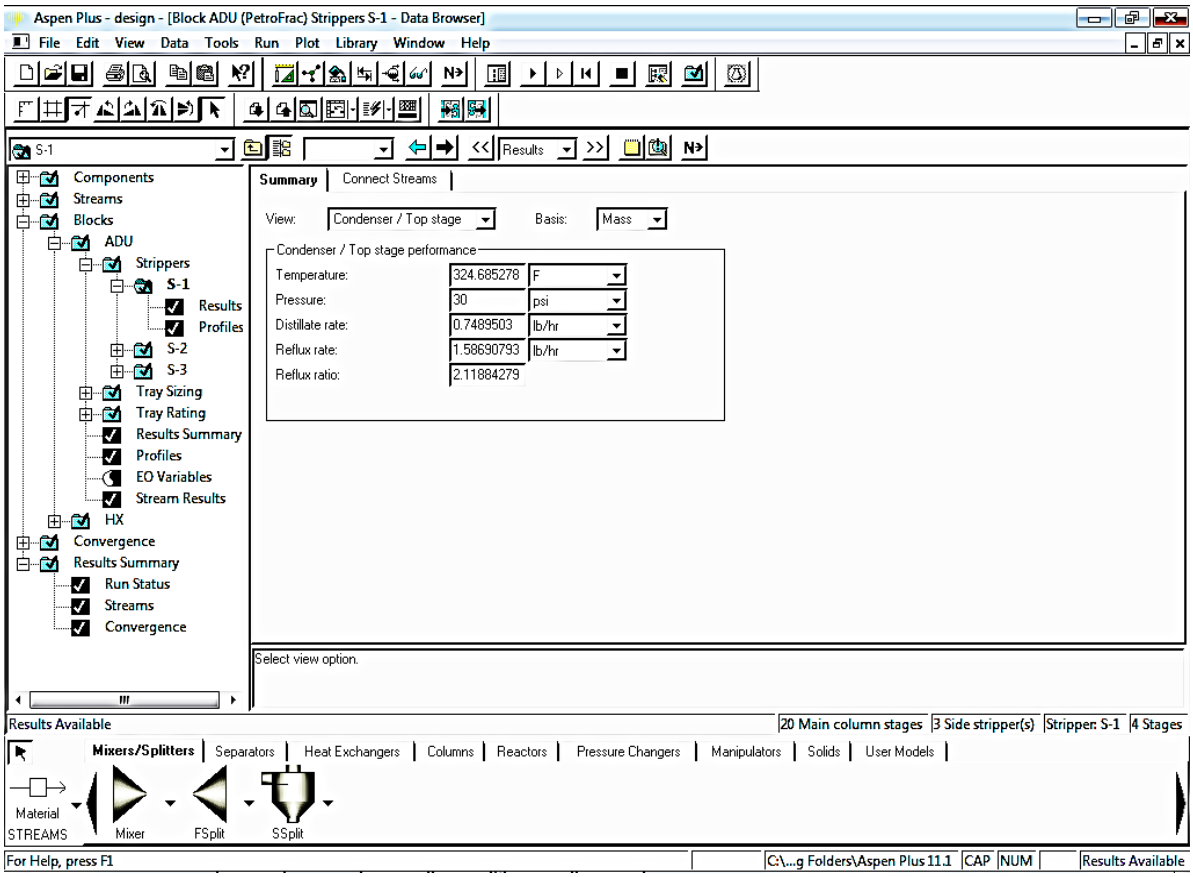


Figure E.29: Result summary for heavy naphtha stripper

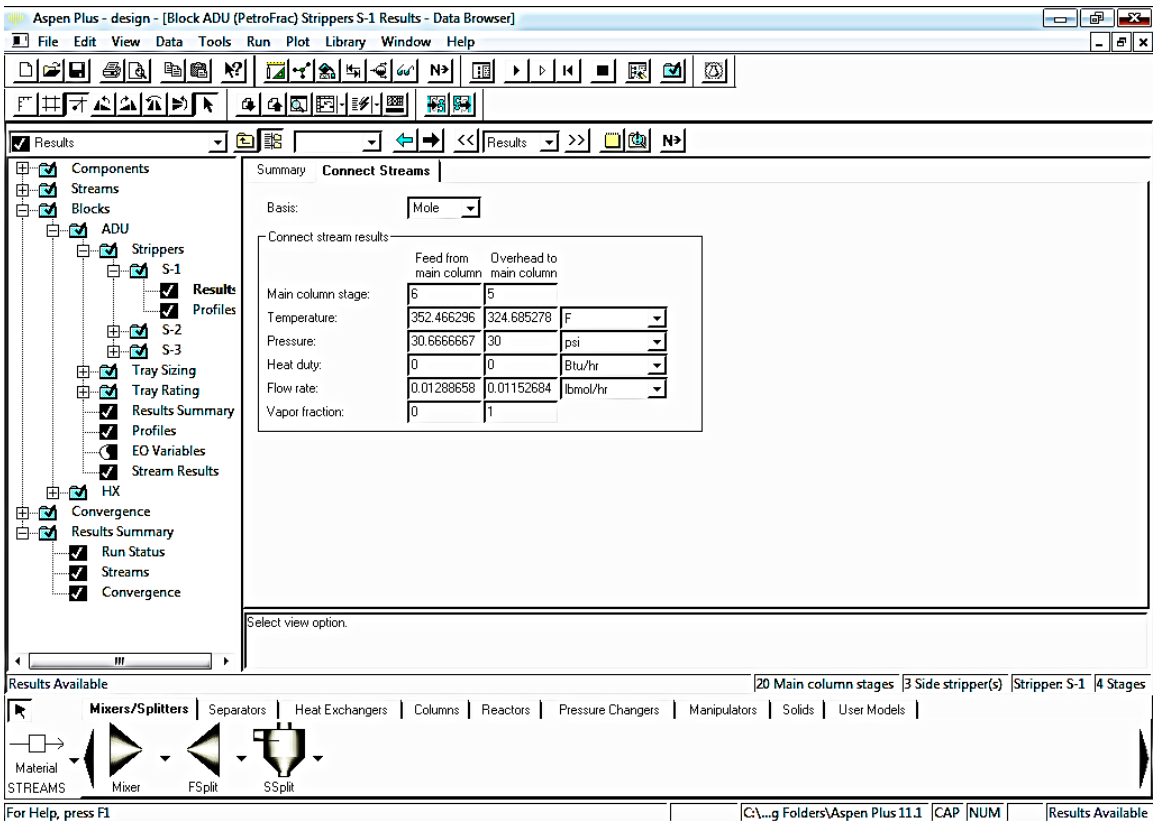


Figure E.30: Connect stream result for heavy naphtha stripper

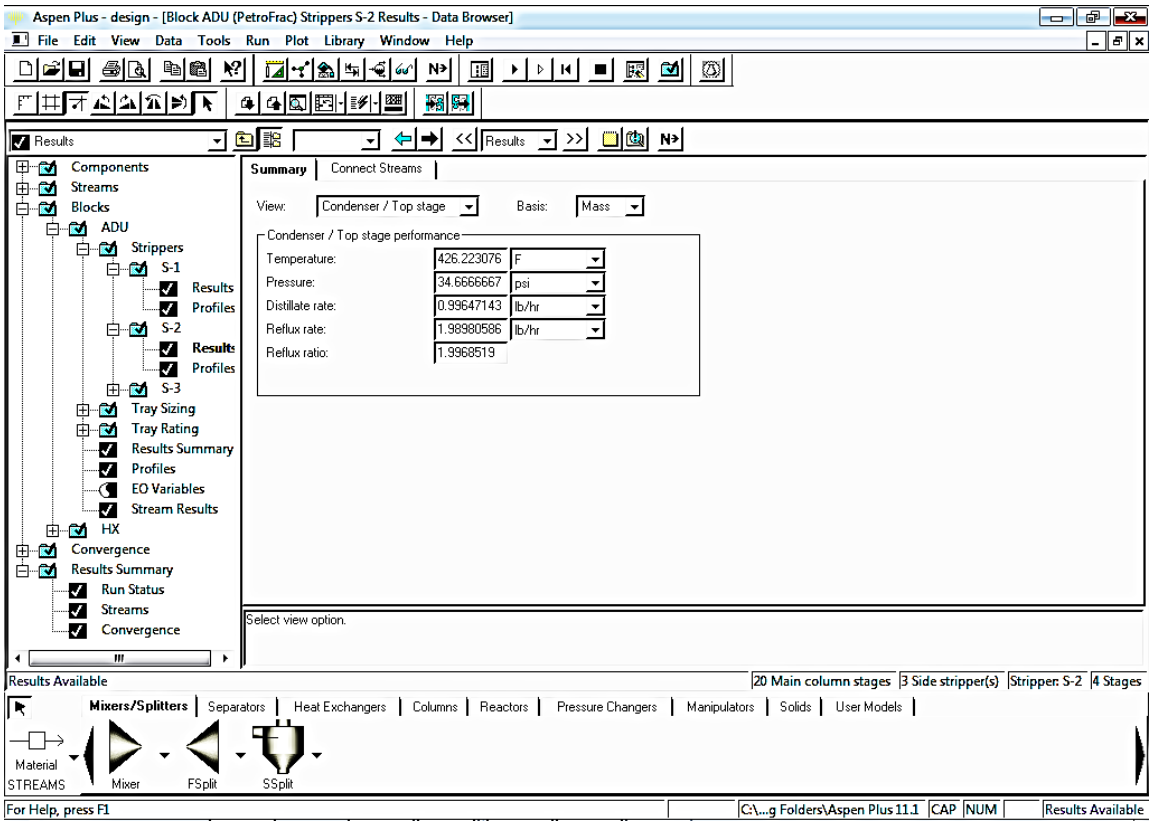


Figure E.31: Result summary for kerosene stripper

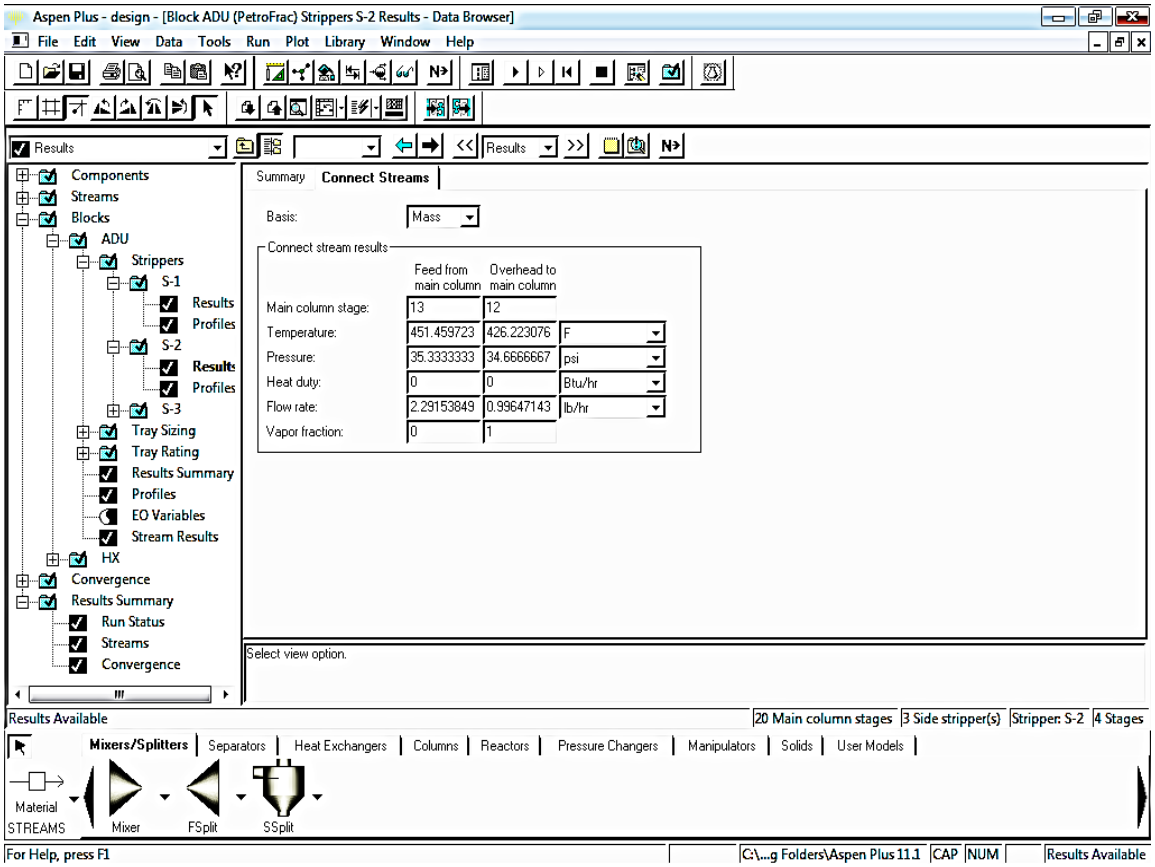


Figure E.32: Connect stream result for kerosene stripper

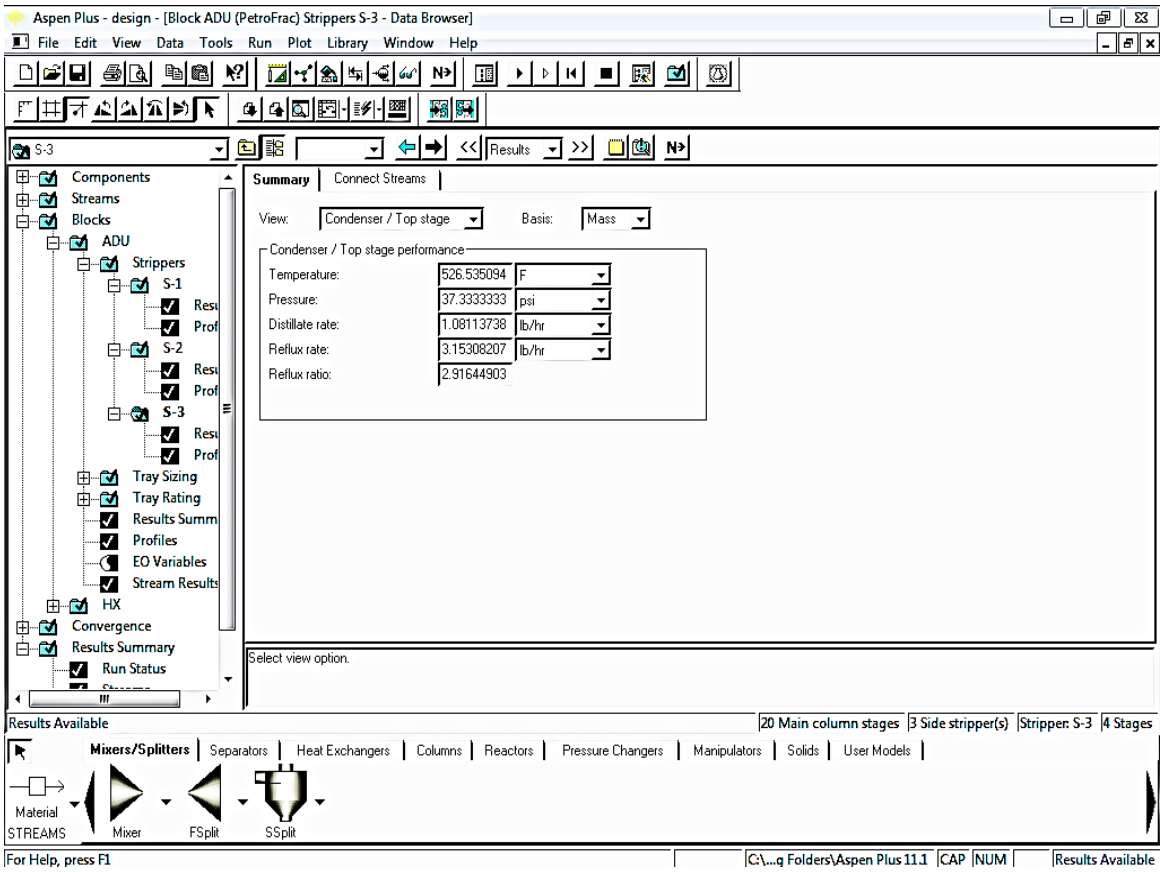


Figure E.33: Result summary for AGO stripper

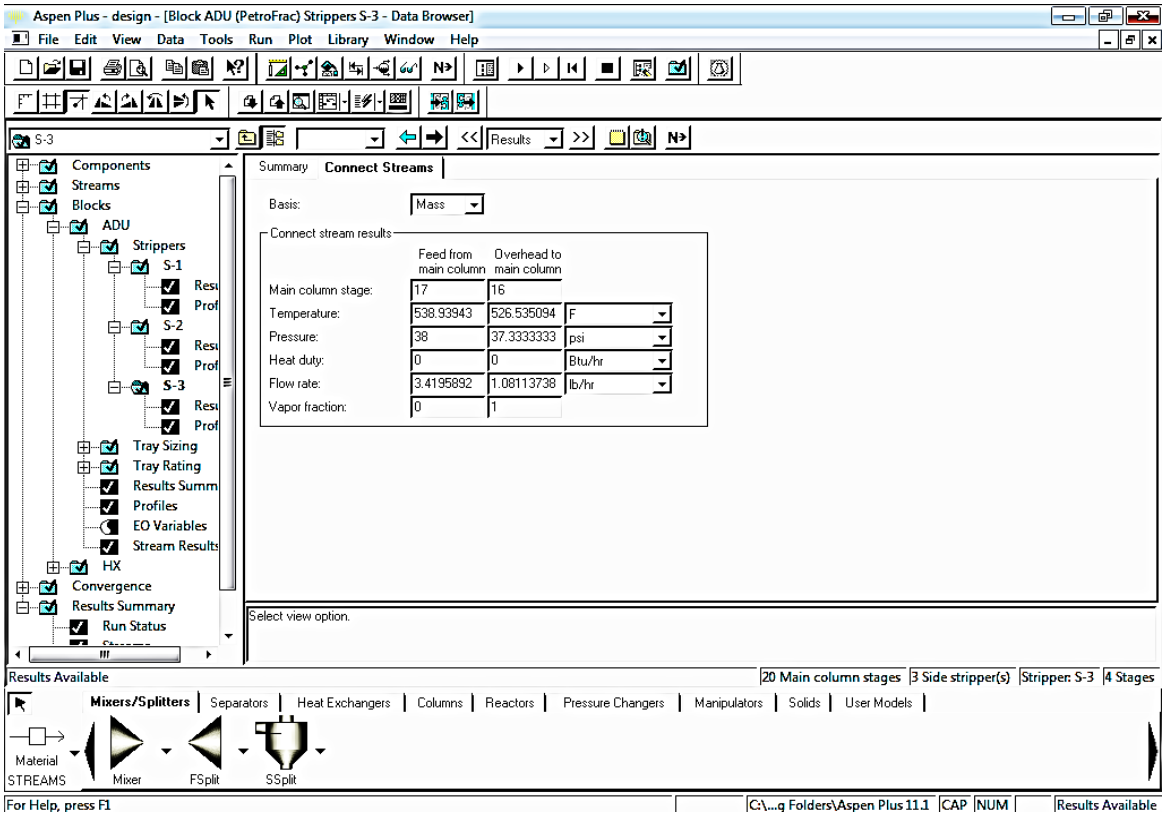


Figure E.34: Connect stream result for AGO stripper

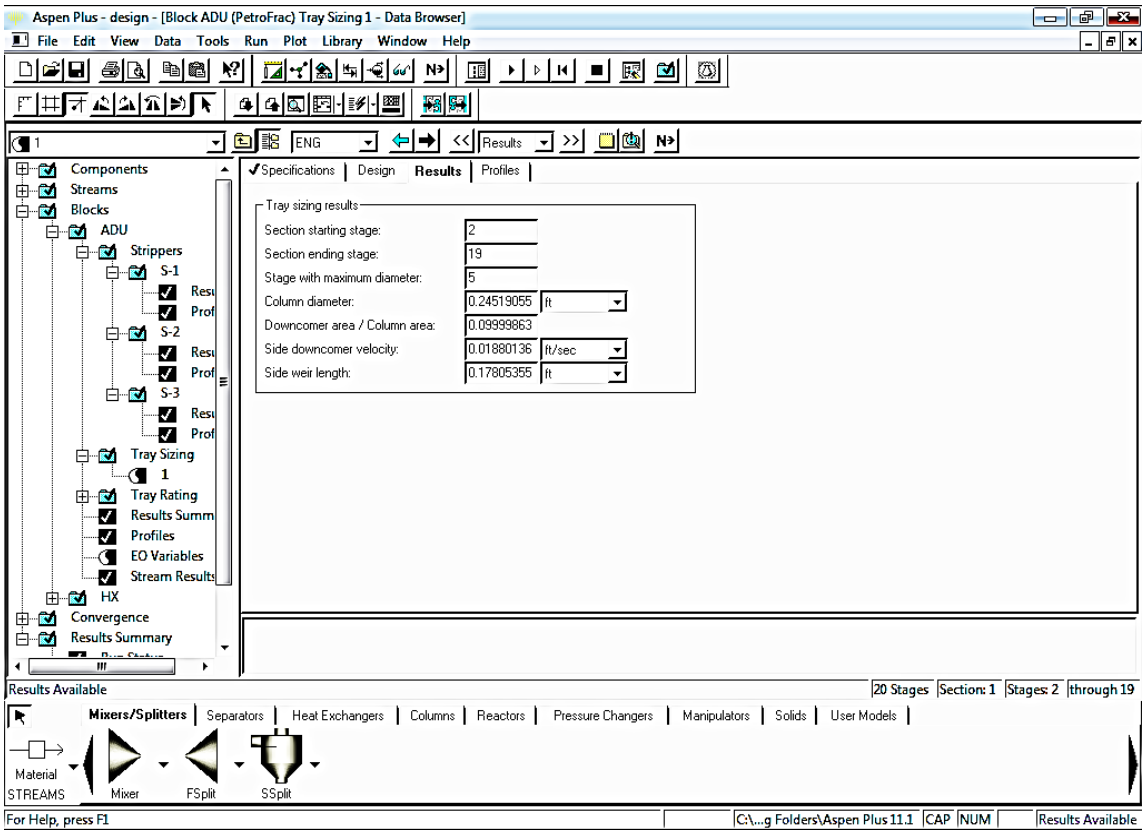


Figure E.35: ADU Tray sizing result

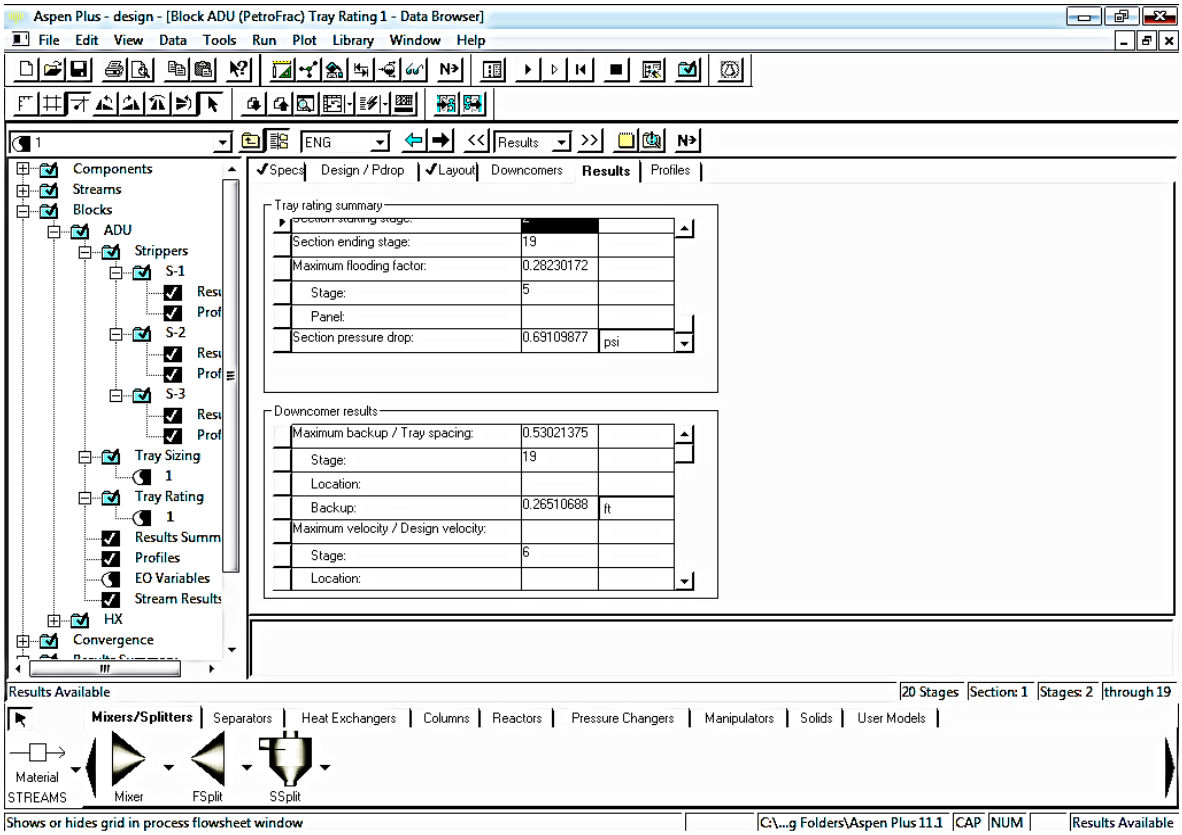


Figure E.36: ADU Tray rating results

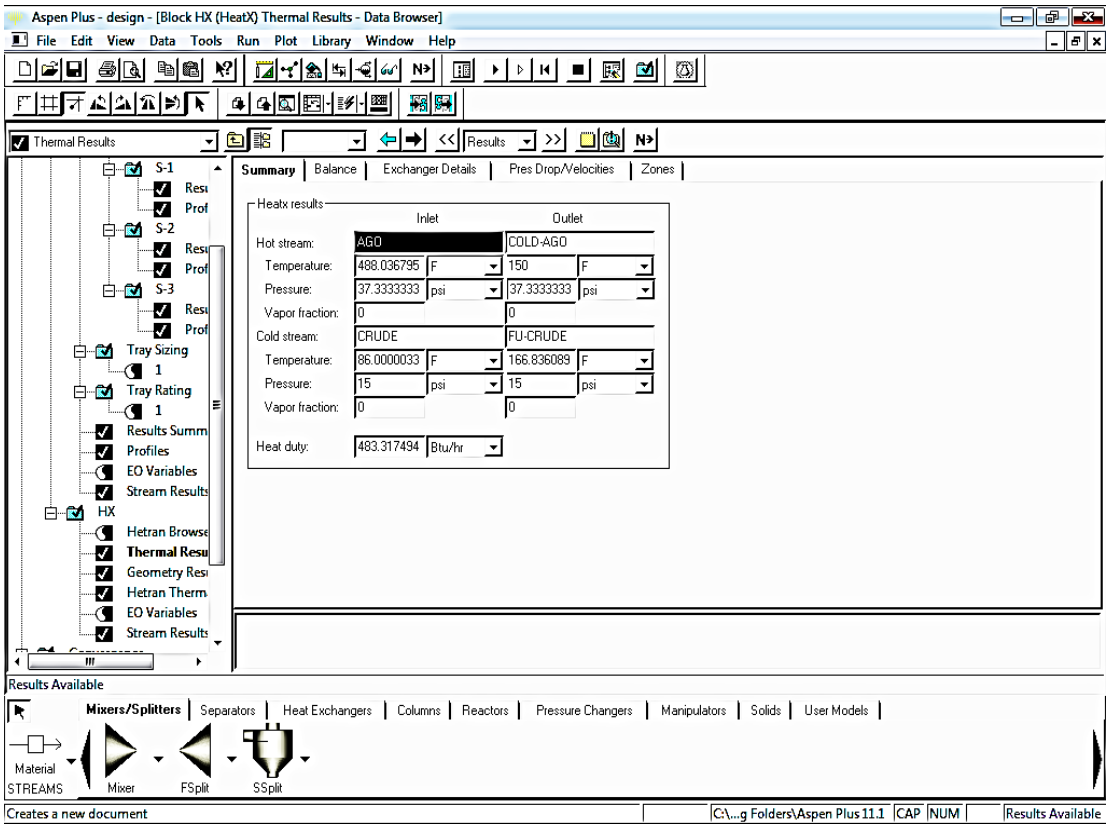


Figure E.37: Summary of heat exchanger result

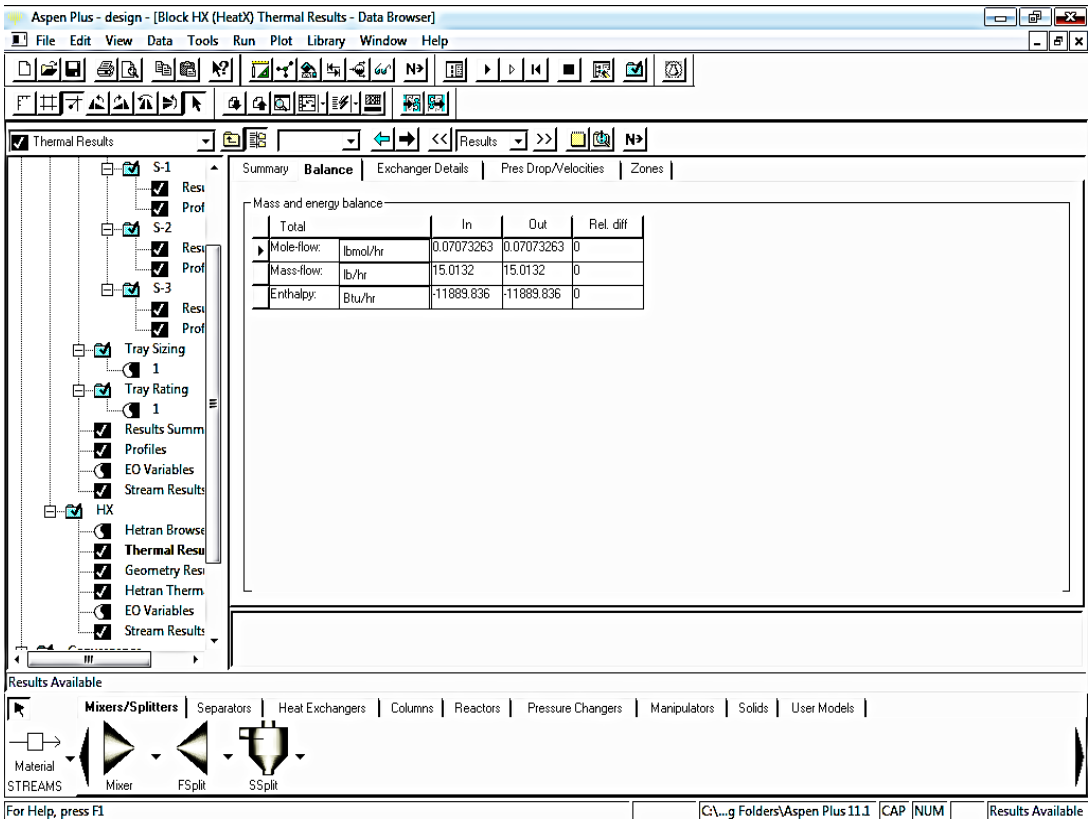


Figure E. 38: Mass and energy balance for heat exchanger

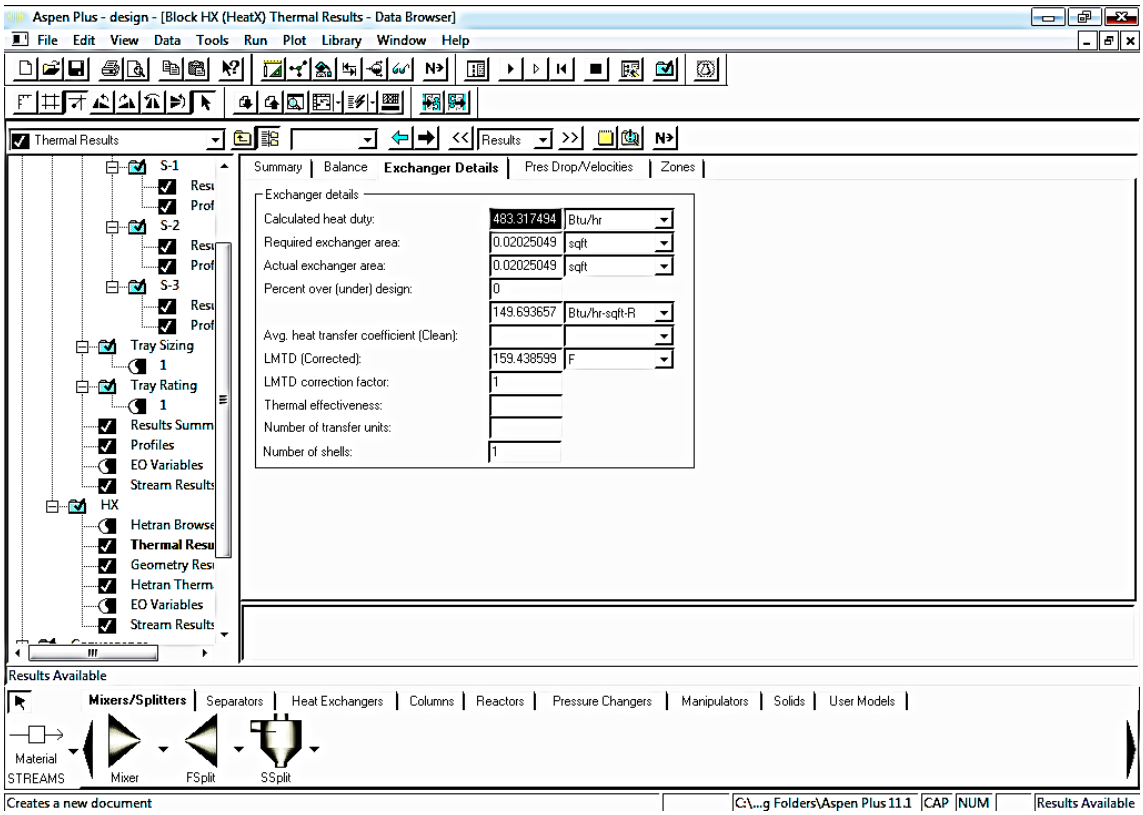


Figure E.39: Exchanger details for heat exchanger

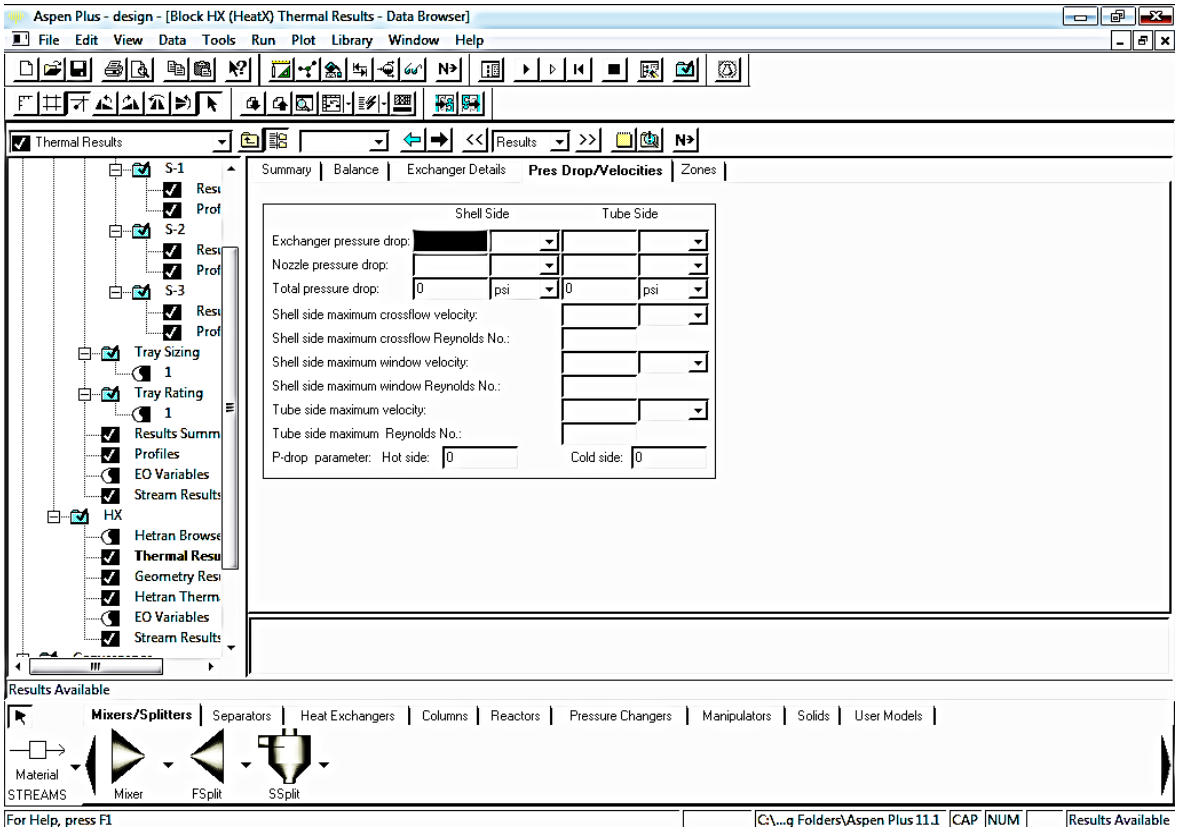
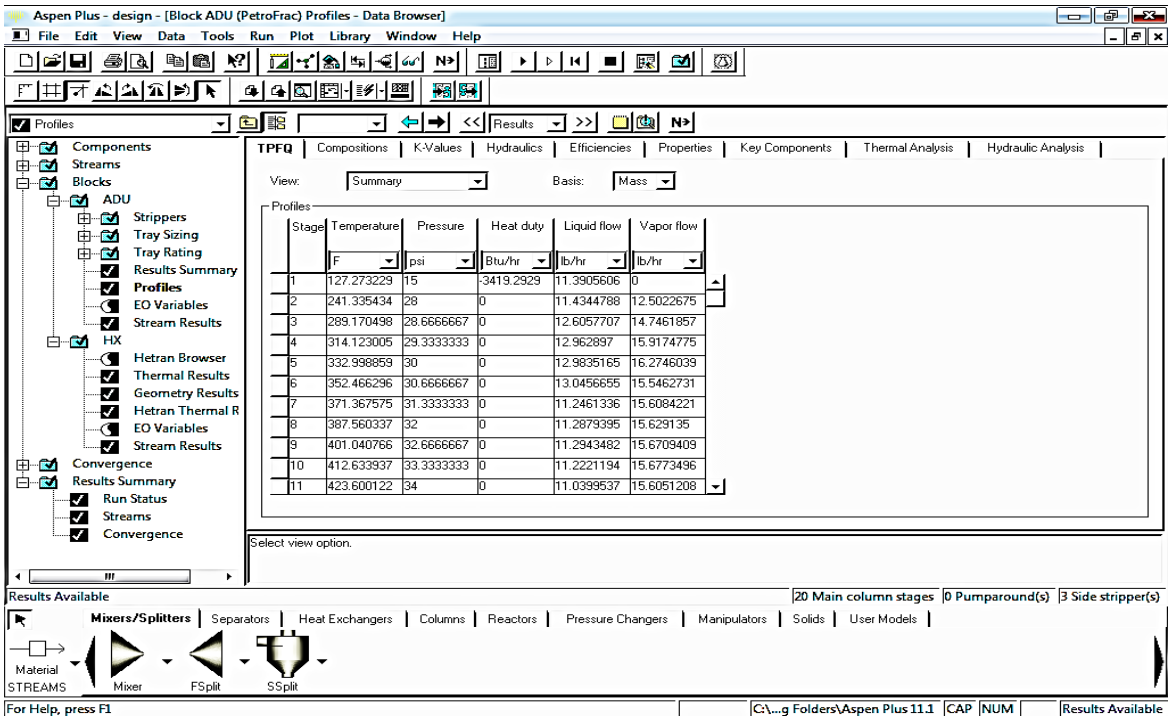
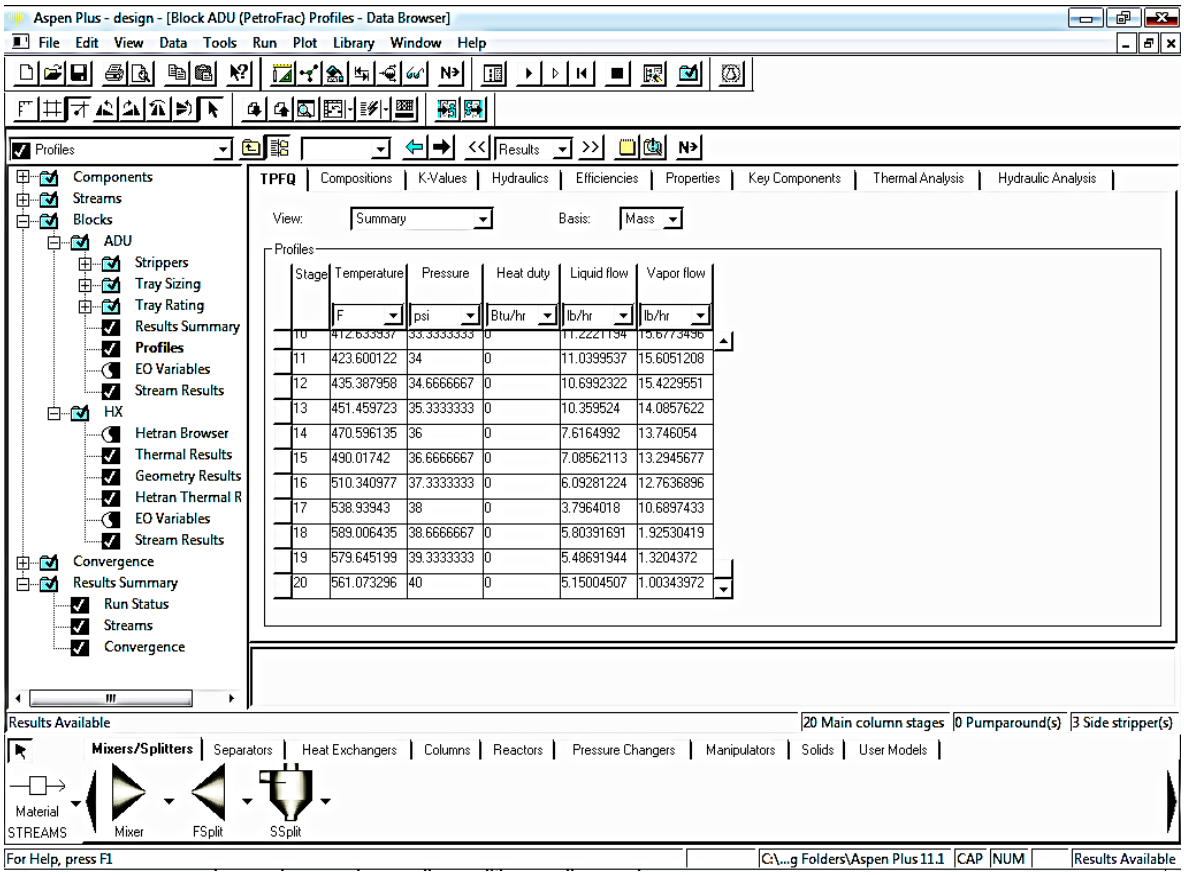


Figure E.40: Pressure drop result for heat exchanger



Stage	Temperature	Pressure	Heat duty	Liquid flow	Vapor flow
	F	psi	Btu/hr	lb/hr	lb/hr
1	127.273229	15	-3419.2929	11.3905606	0
2	241.335434	28	0	11.4344788	12.5022675
3	289.170498	28.6666667	0	12.6057707	14.7461857
4	314.123005	29.3333333	0	12.962897	15.9174775
5	332.998859	30	0	12.9835165	16.2746039
6	352.466296	30.6666667	0	13.0456655	15.5462731
7	371.367575	31.3333333	0	11.2461336	15.6084221
8	387.560337	32	0	11.2879395	15.629135
9	401.040766	32.6666667	0	11.2943482	15.6709409
10	412.633937	33.3333333	0	11.2221194	15.6773496
11	423.600122	34	0	11.0399537	15.6051208

Figure E.41: ADU Profile results



Stage	Temperature	Pressure	Heat duty	Liquid flow	Vapor flow
	F	psi	Btu/hr	lb/hr	lb/hr
10	412.633937	33.3333333	0	11.2221194	15.6773496
11	423.600122	34	0	11.0399537	15.6051208
12	435.387958	34.6666667	0	10.6992322	15.4229551
13	451.459723	35.3333333	0	10.359524	14.0857622
14	470.596135	36	0	7.6164992	13.746054
15	490.01742	36.6666667	0	7.08562113	13.2945677
16	510.340977	37.3333333	0	6.09281224	12.7636896
17	538.93943	38	0	3.7964018	10.6897433
18	589.006435	38.6666667	0	5.80391691	1.92530419
19	579.645199	39.3333333	0	5.48691944	1.3204372
20	561.073296	40	0	5.15004507	1.00343972

Figure E.42: ADU Profile results continues

APPENDIX F

F.0 Aspen Plus Result for Vacuum Distillation Unit

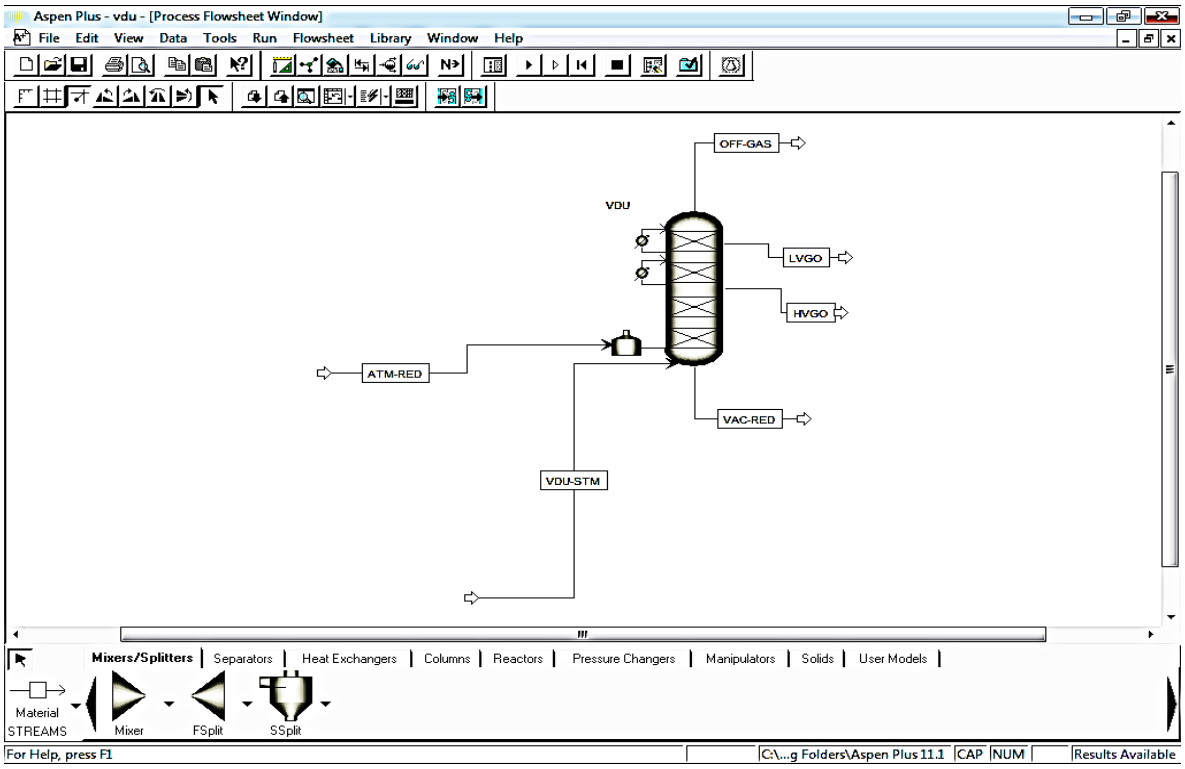


Figure F.1: Flowsheet for VDU

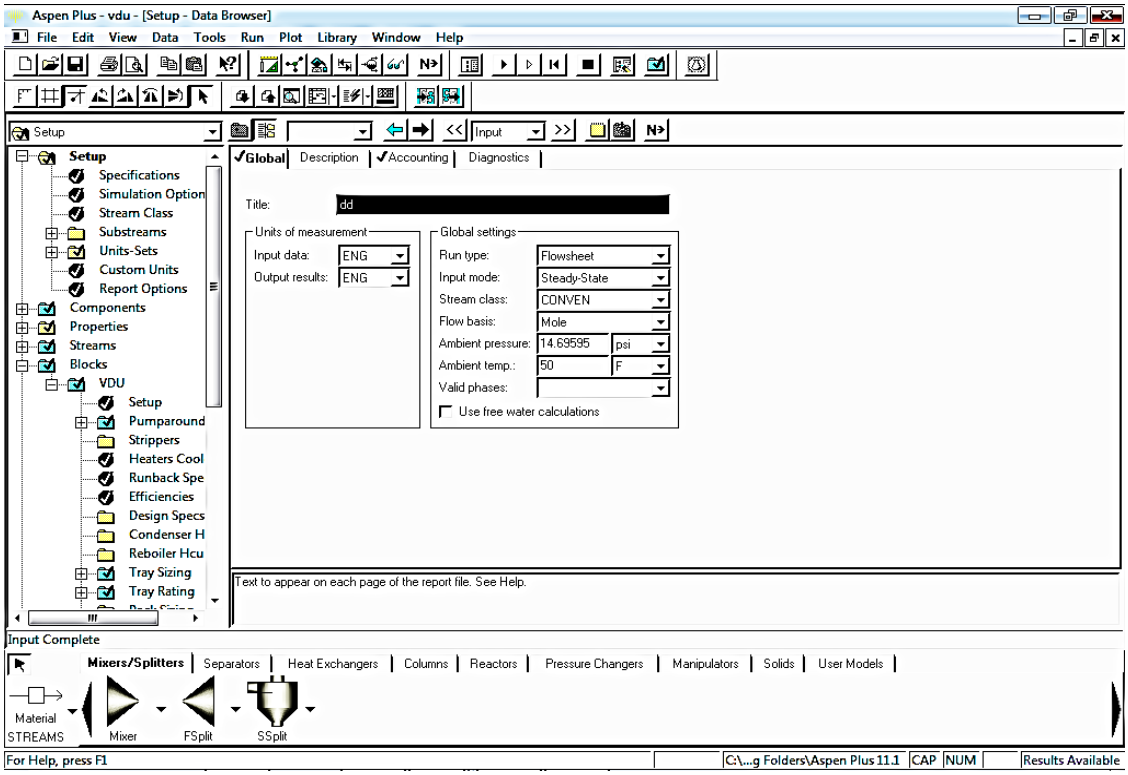


Figure F.2: Setup specification for VDU

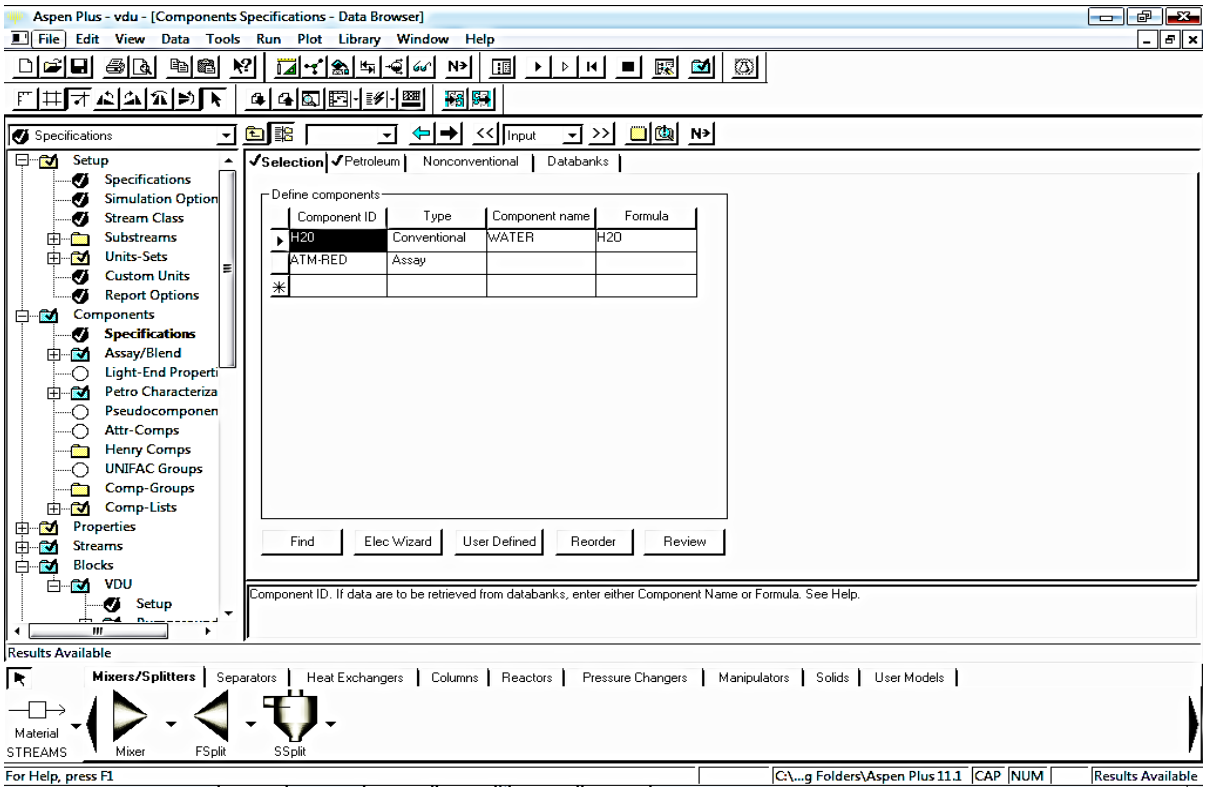


Figure F.3: component specification for VDU

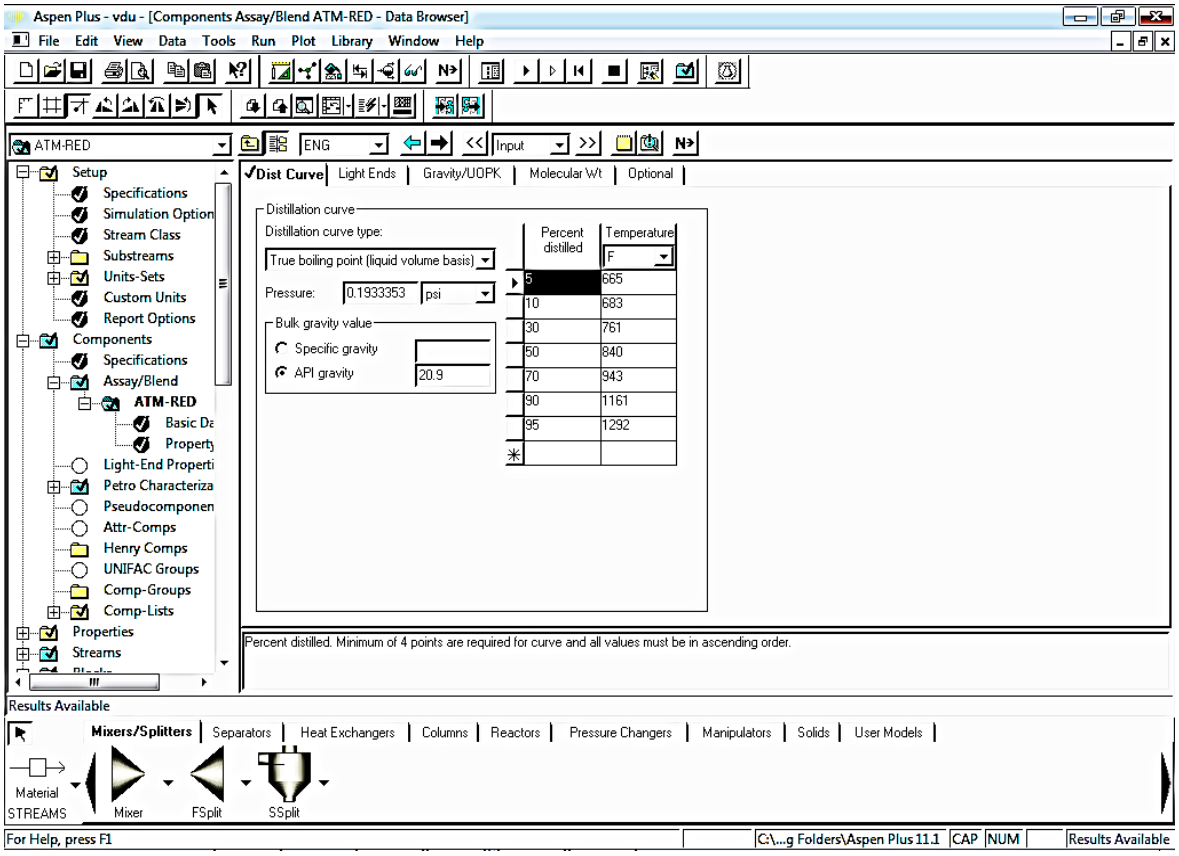


Figure F.4: Assay specification for Atmospheric residue

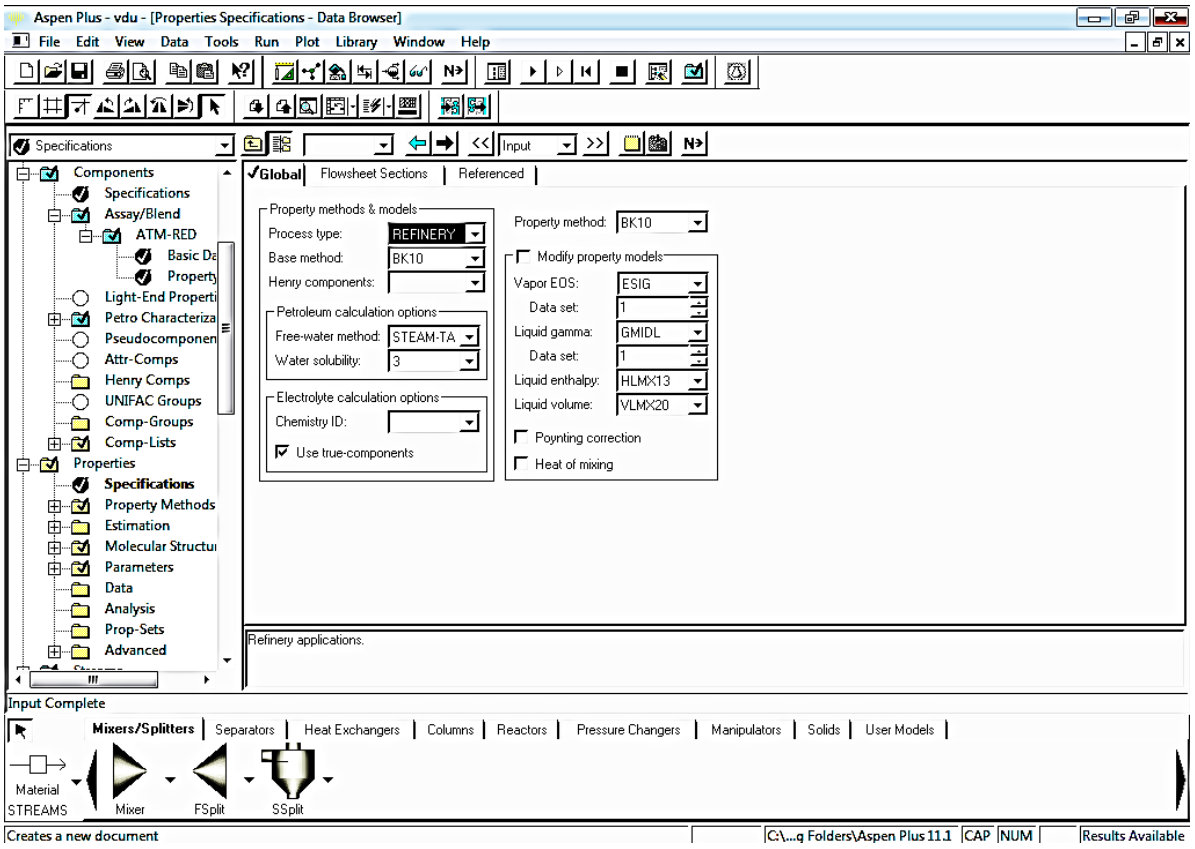


Figure F.5: property specification for VDU

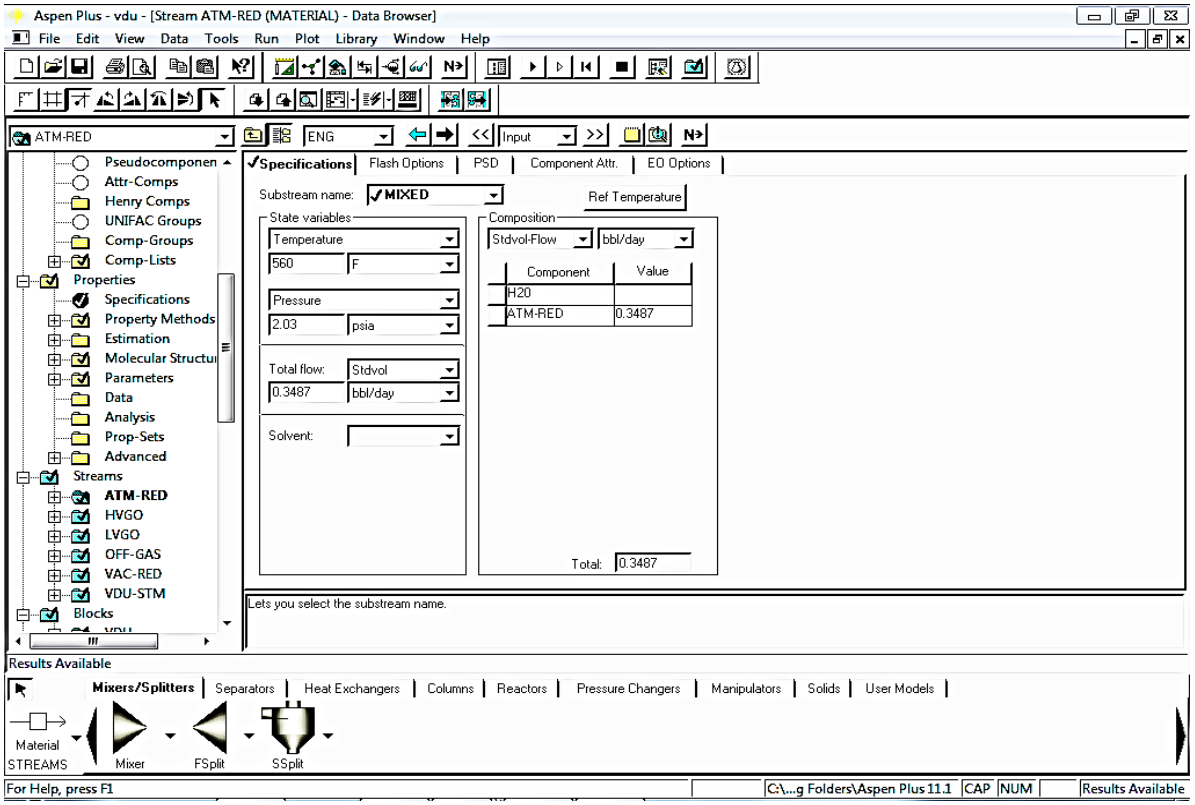


Figure F.6: Specification of Atm residue stream to VDU

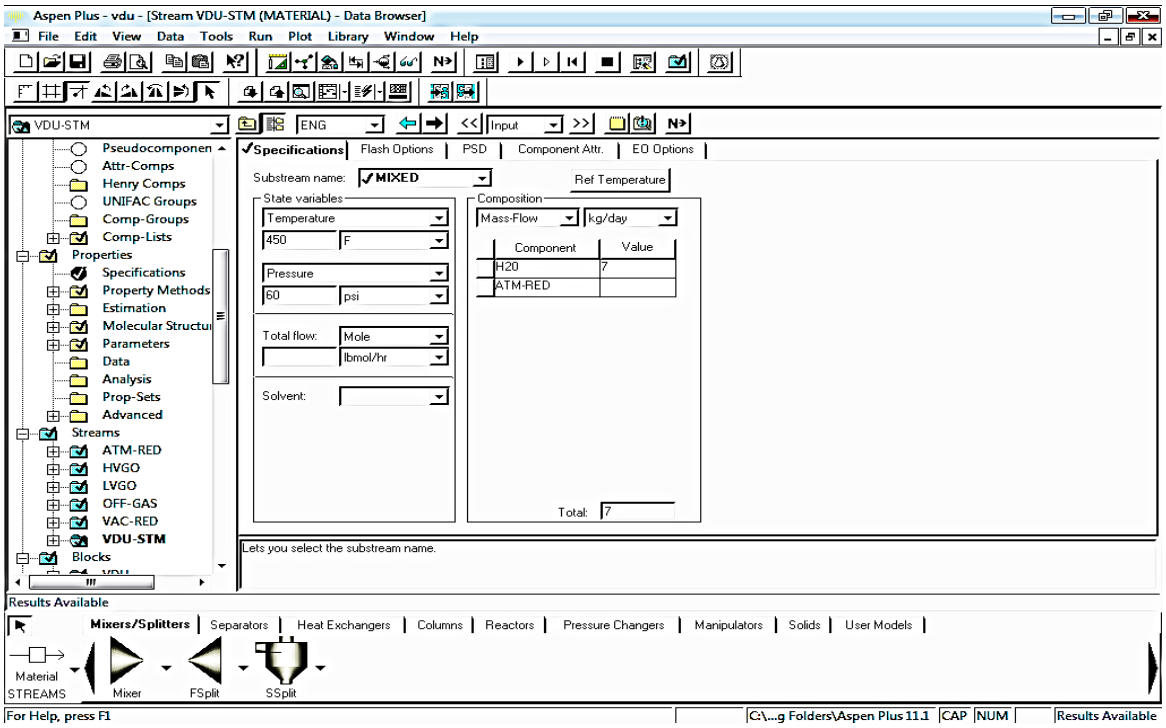


Figure F.7: Specification of Steam stream to

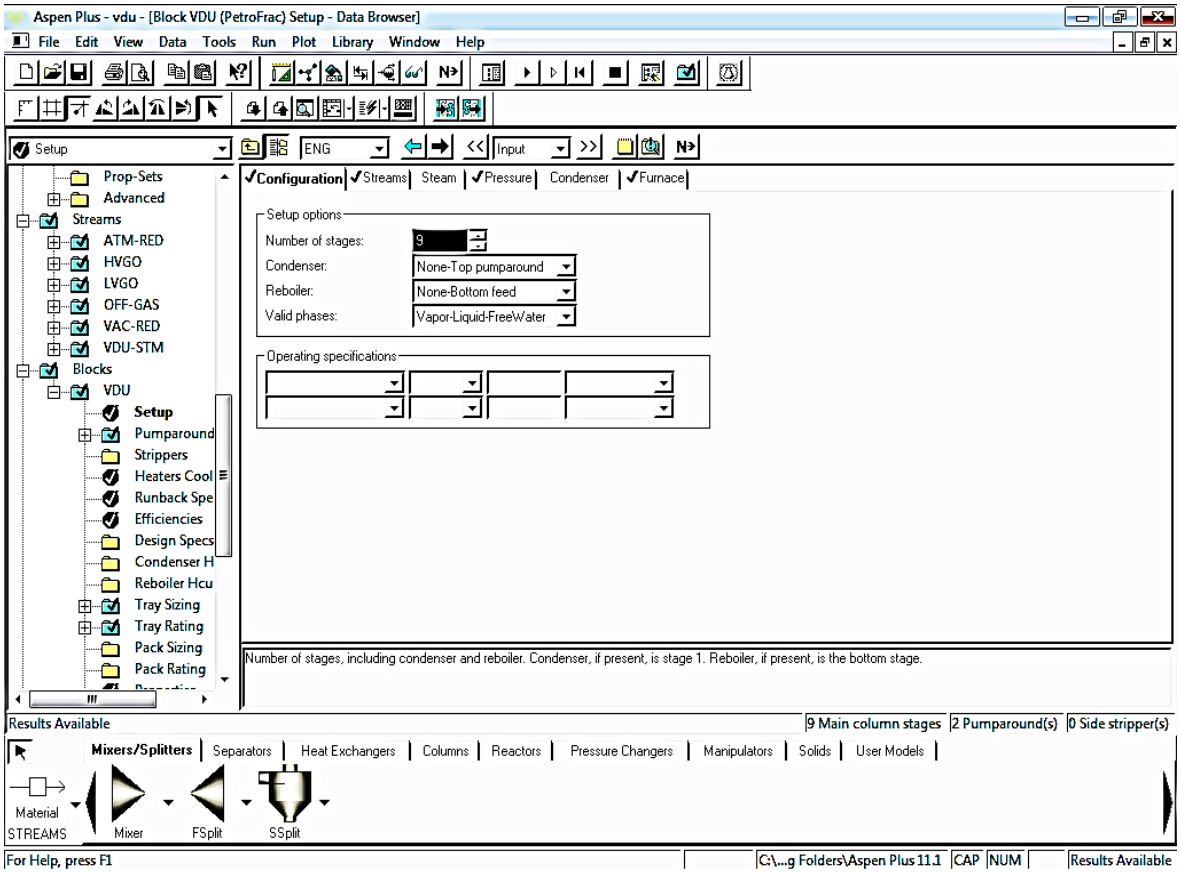


Figure F.8: VDU configuration specification

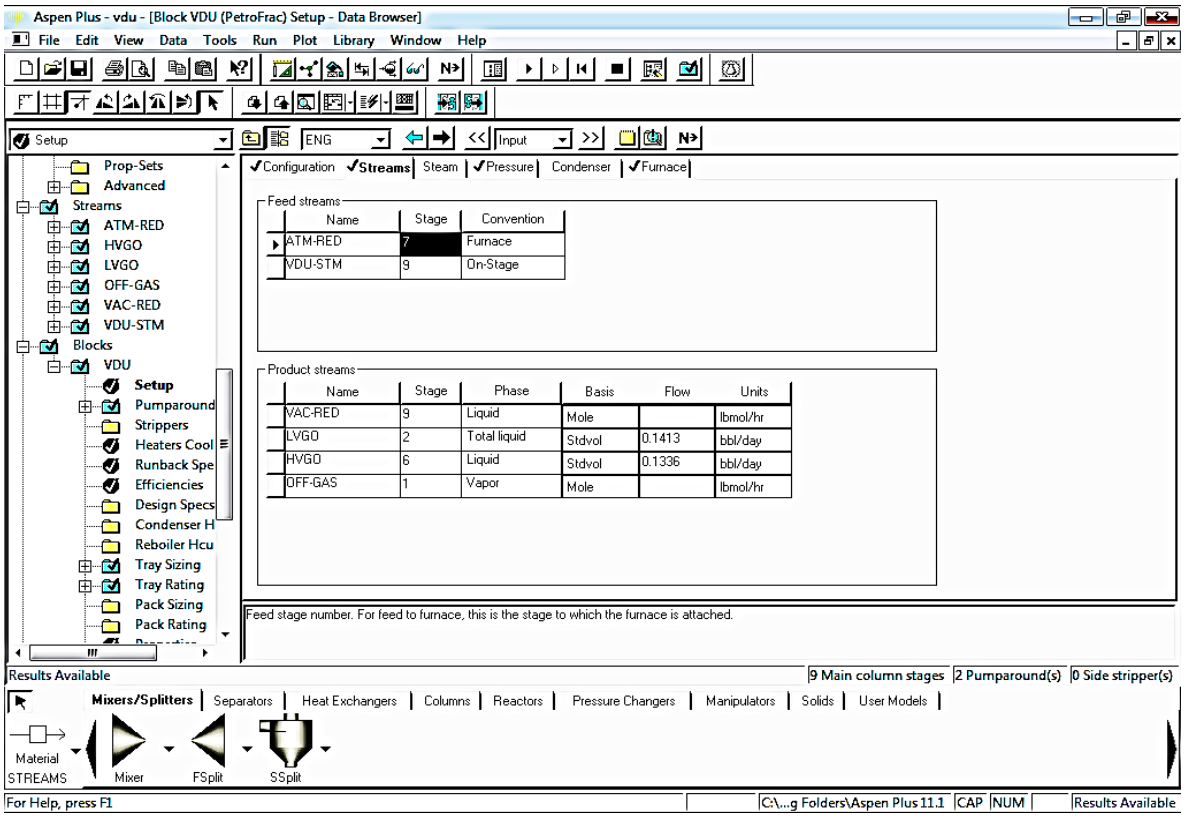


Figure F.9: VDU stream specifications

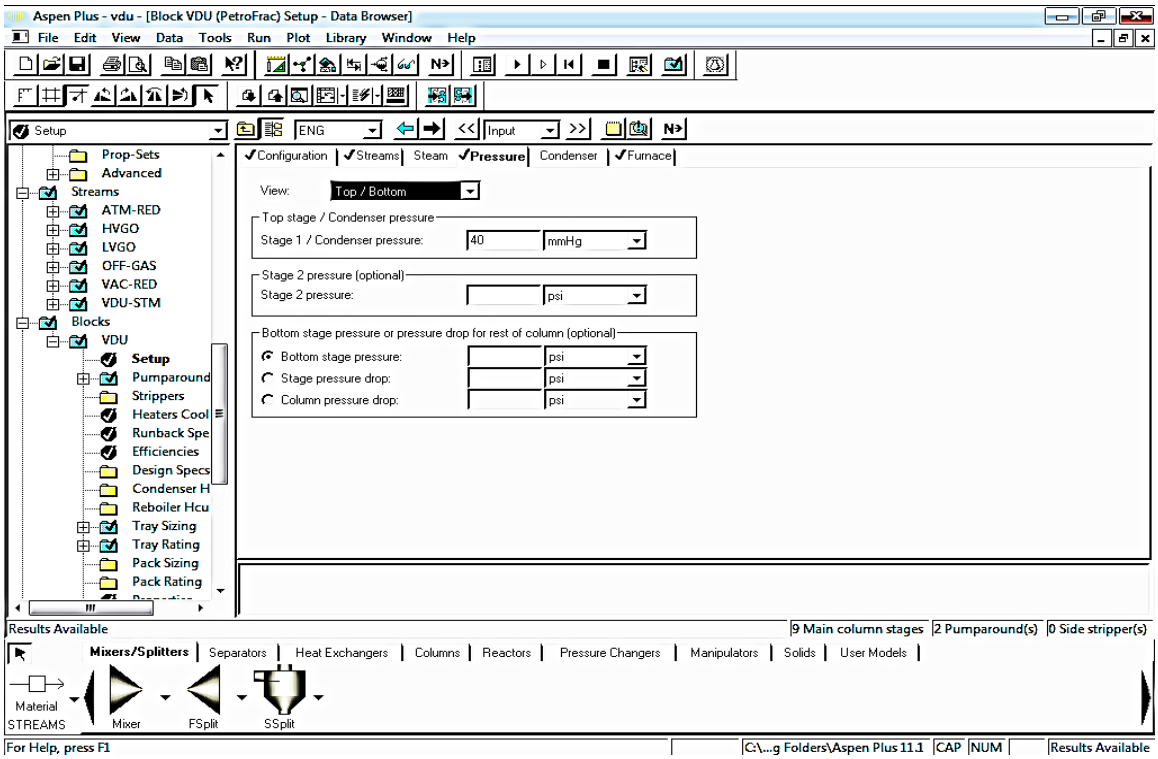


Figure F.10: VDU pressure specification

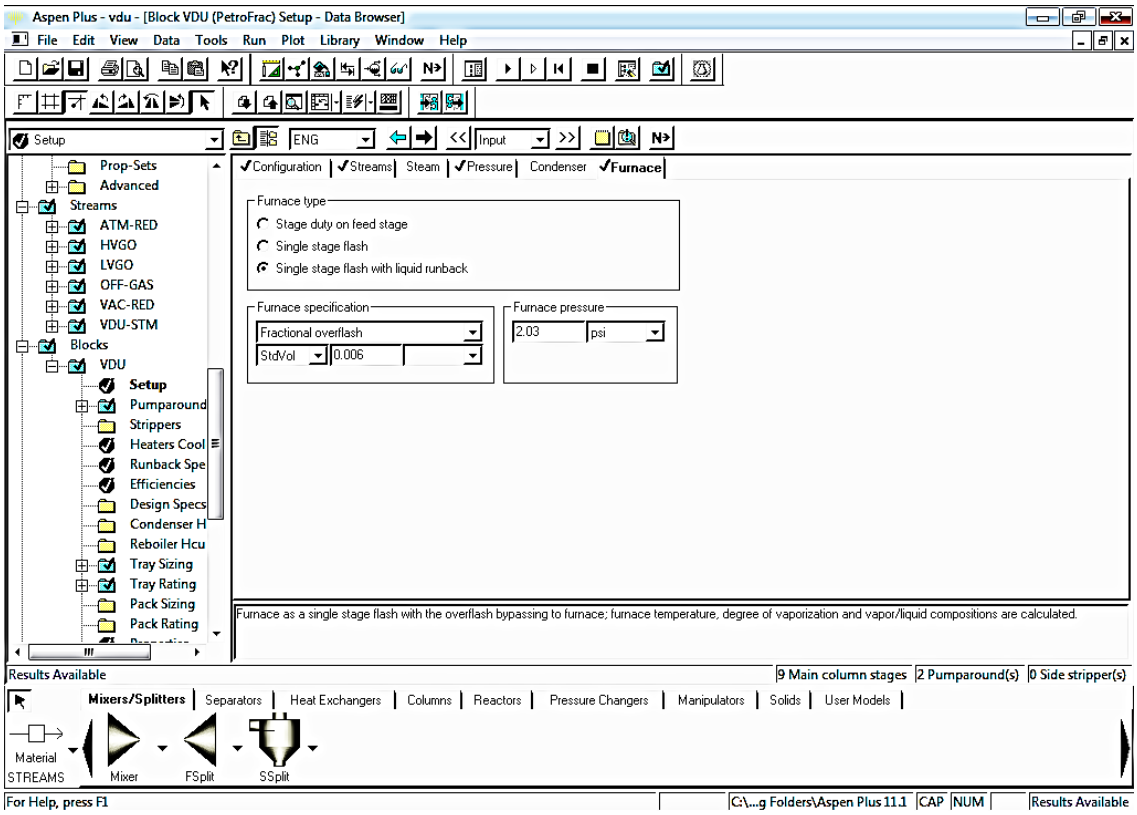


Figure F.11: VDU furnace specificatio

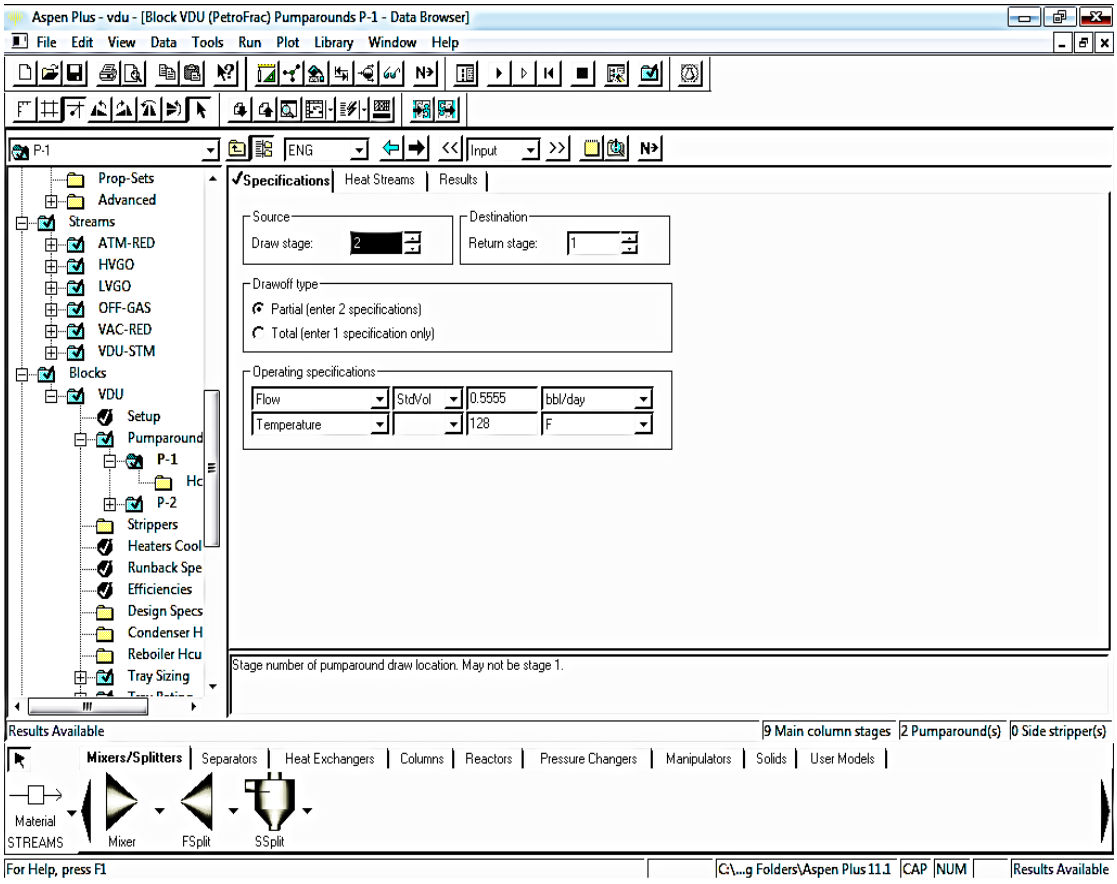


Figure F.12: VDU Top pump around specification

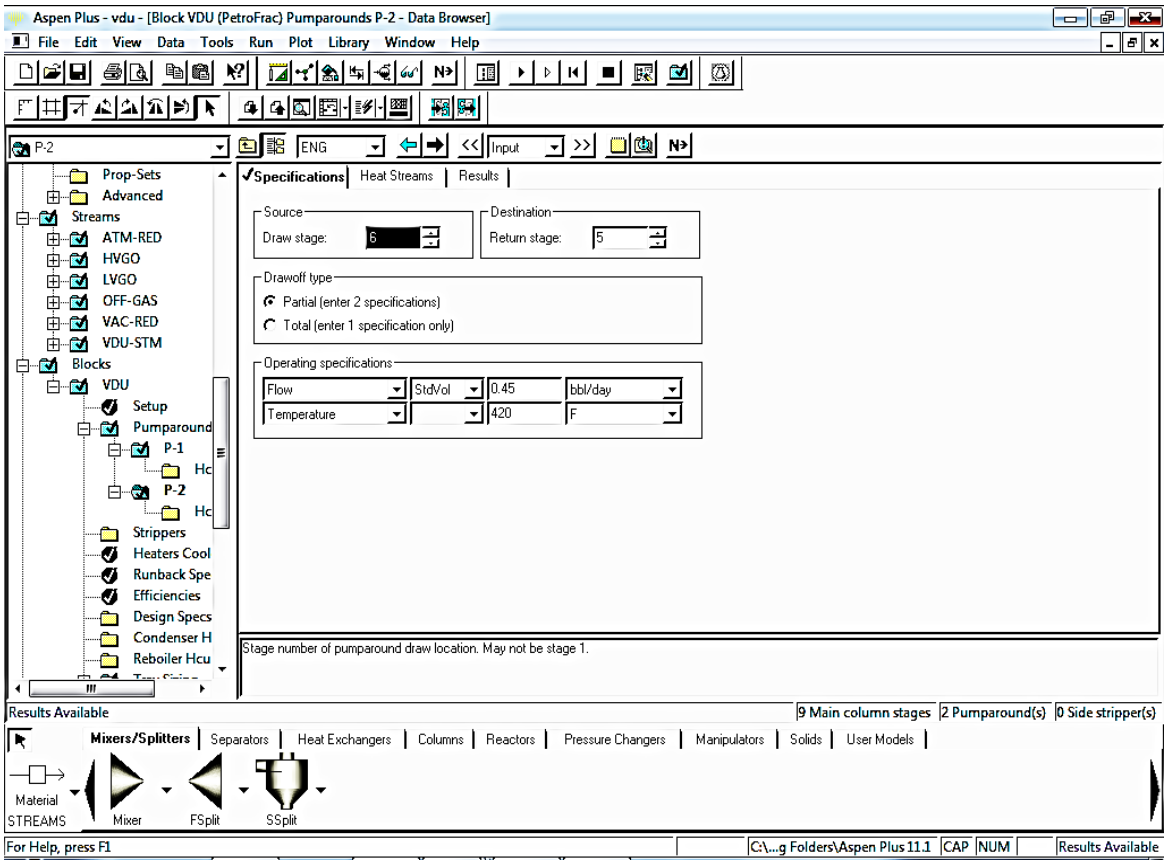


Figure F.13: VDU bottom pump around specificatio

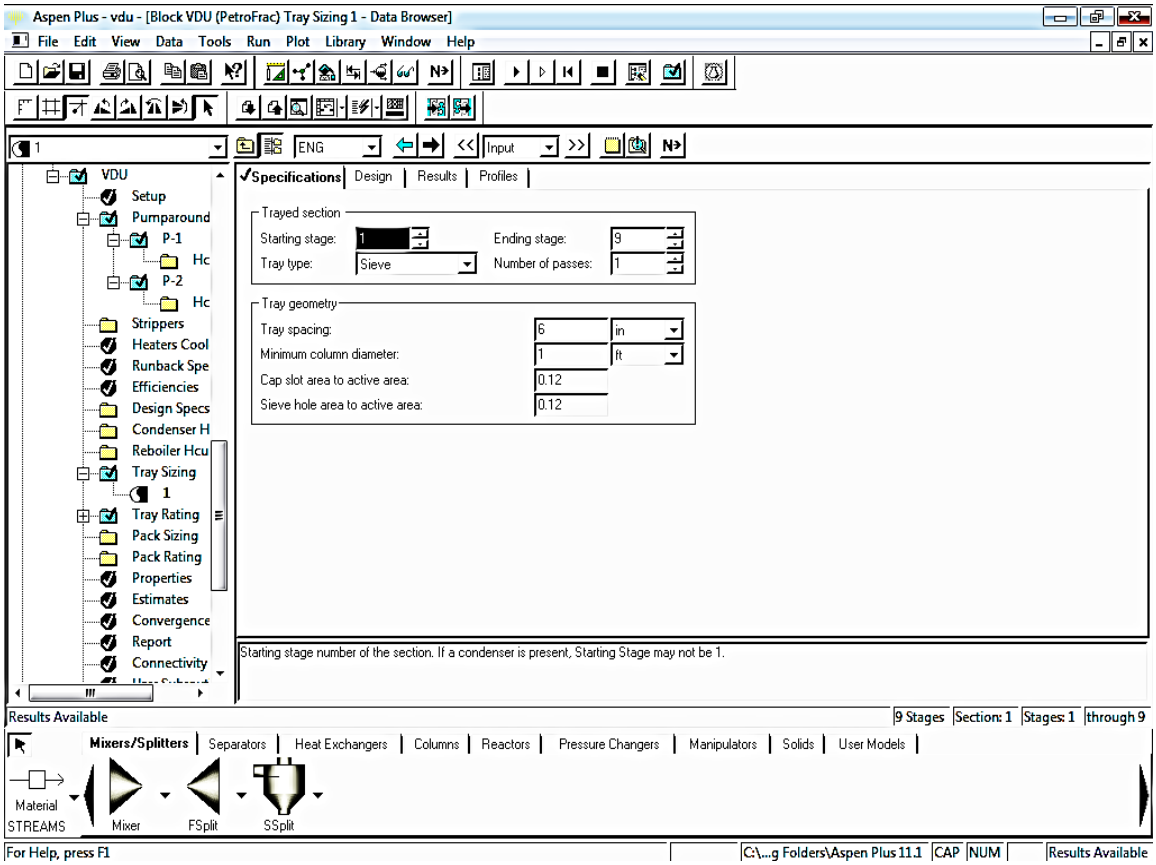


Figure F.14: VDU tray sizing specifications

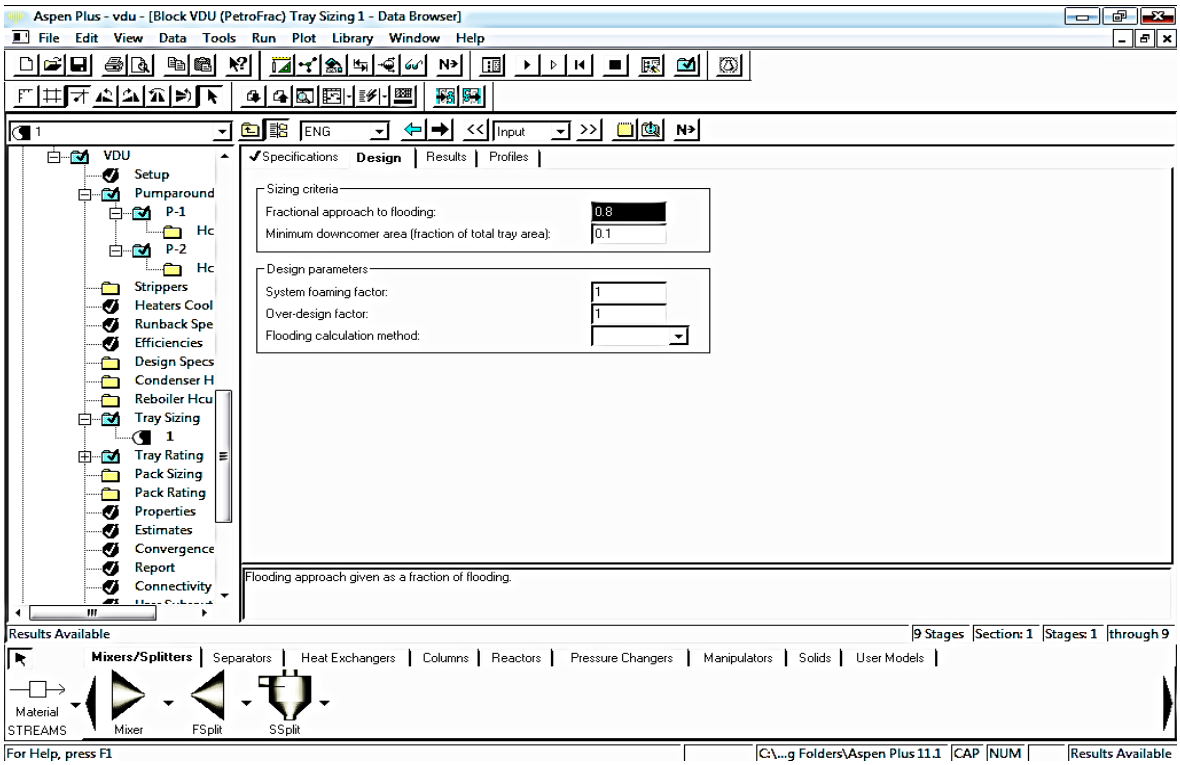


Figure F.15: VDU tray sizing design specs.

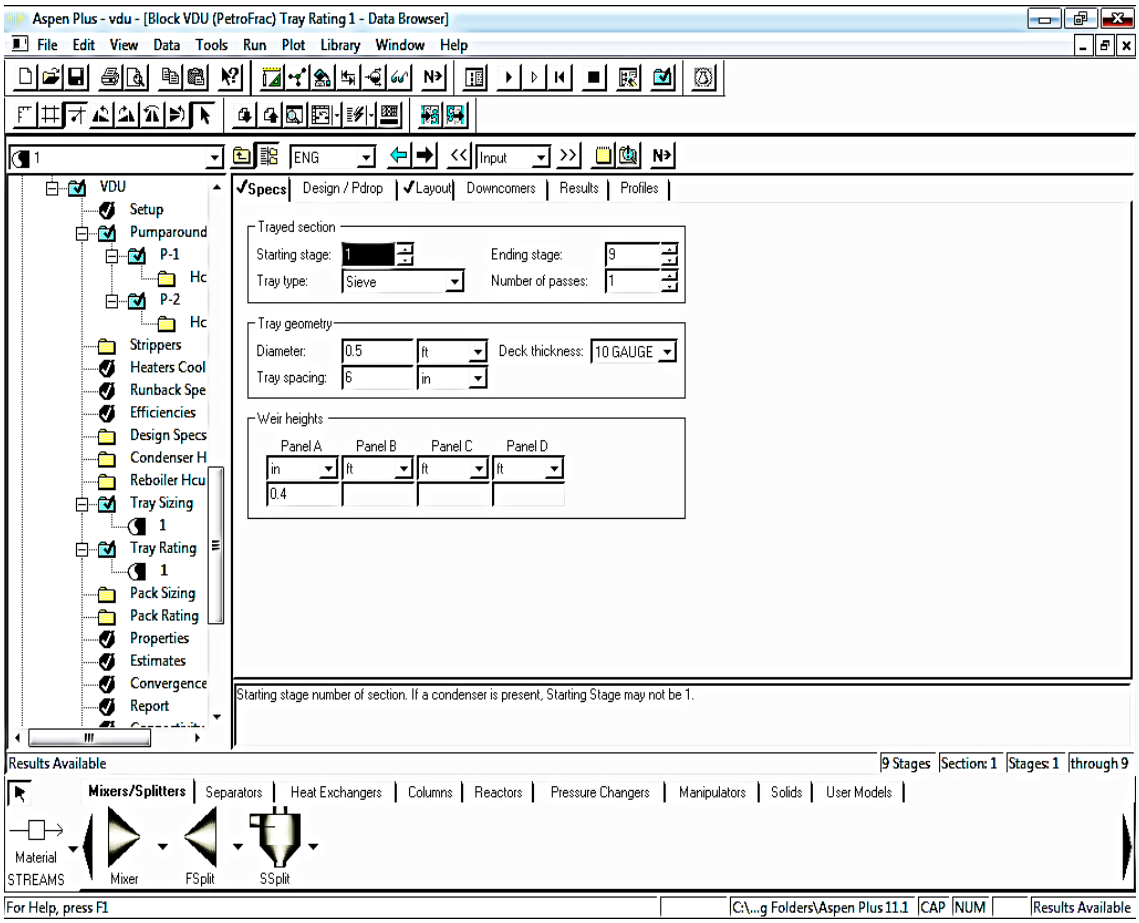


Figure F.16: VDU tray rating spe

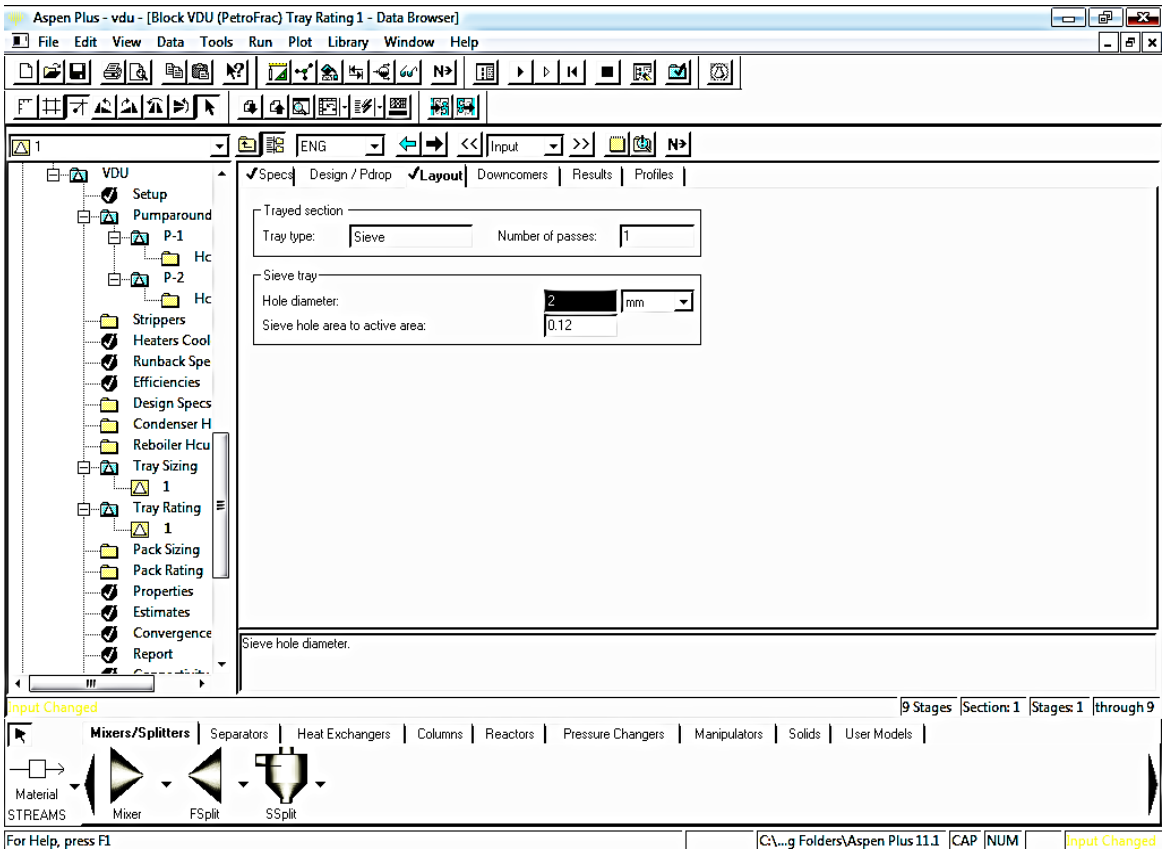


Figure F.17: VDU tray rating layout

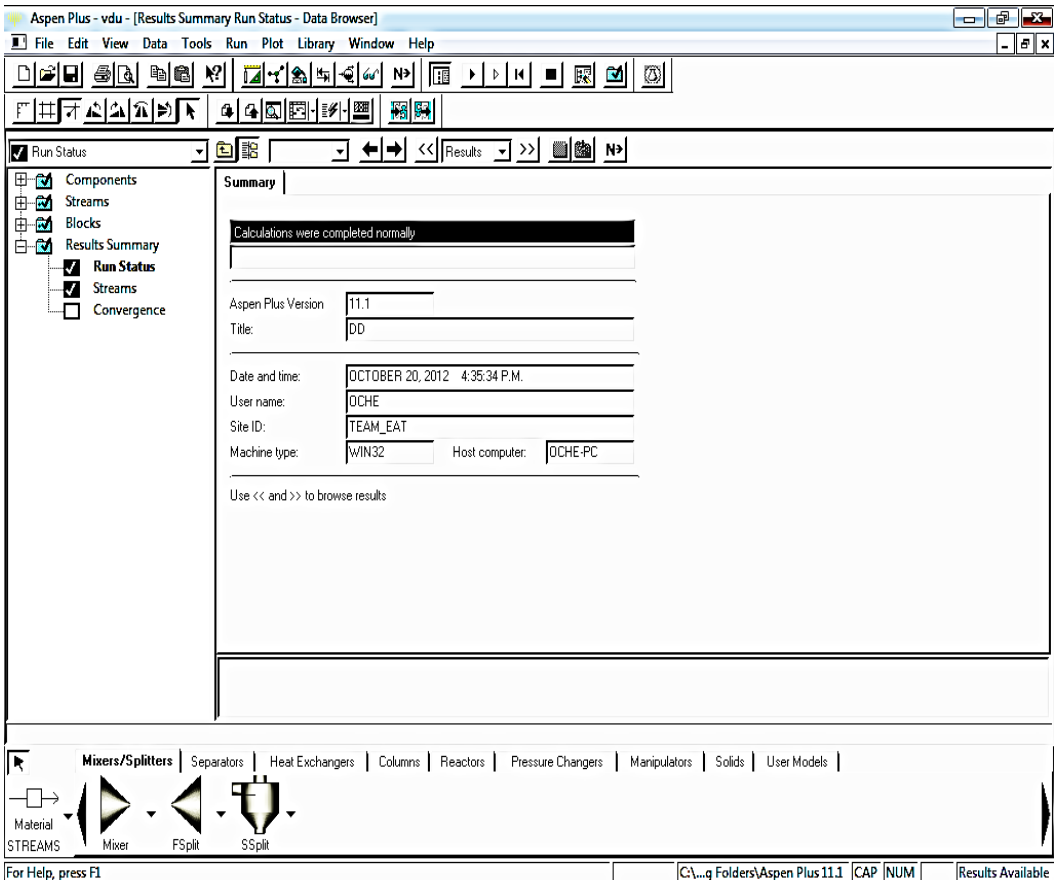


Figure F.18: VDU result run status

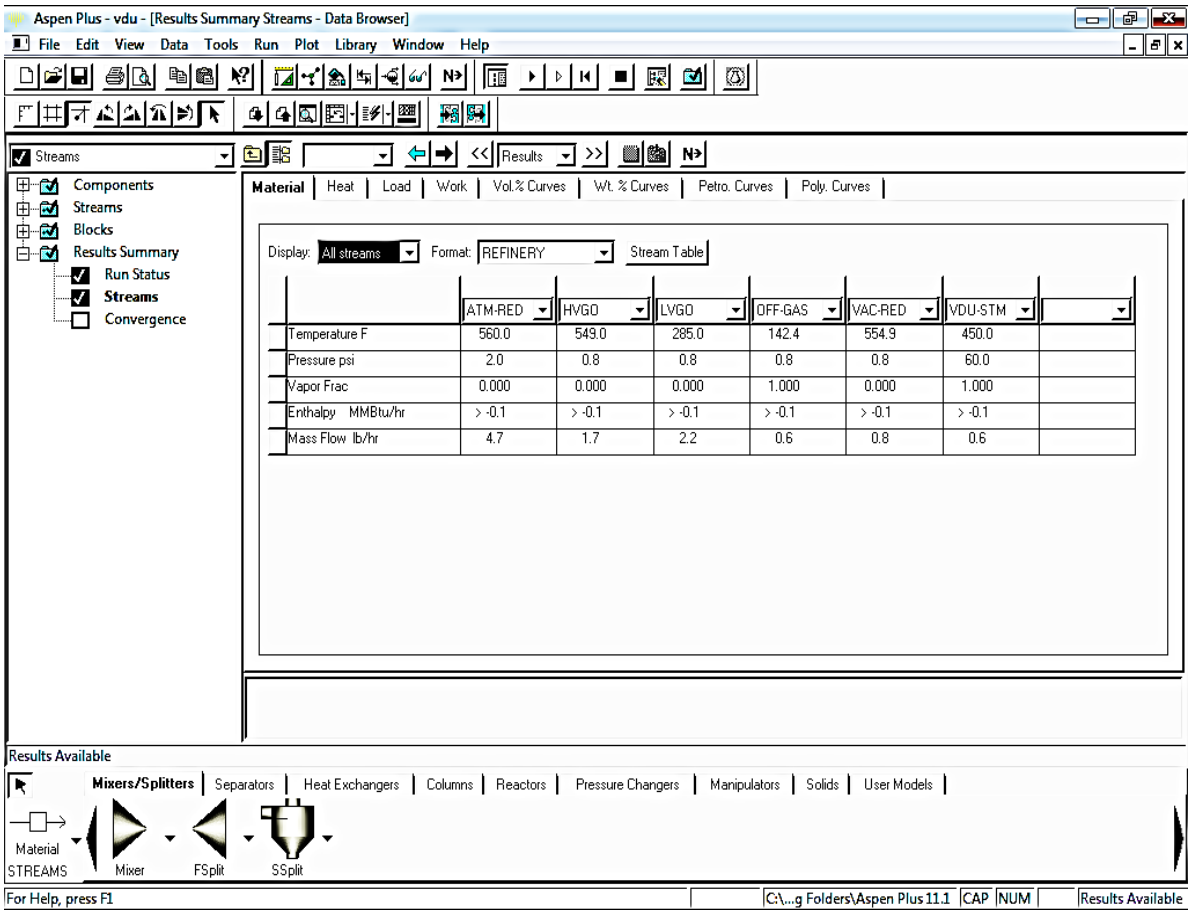


Figure F.19: VDU stream result

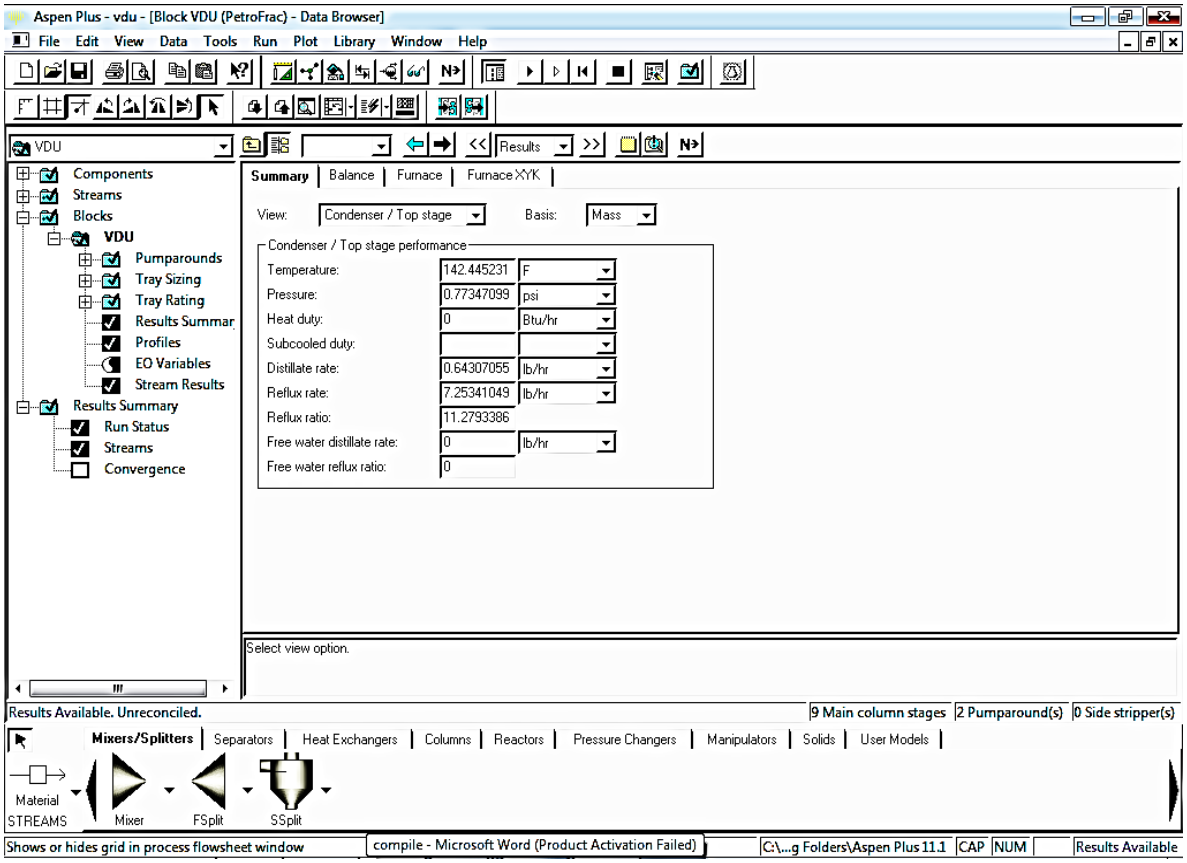


Figure F.20: VDU result summary

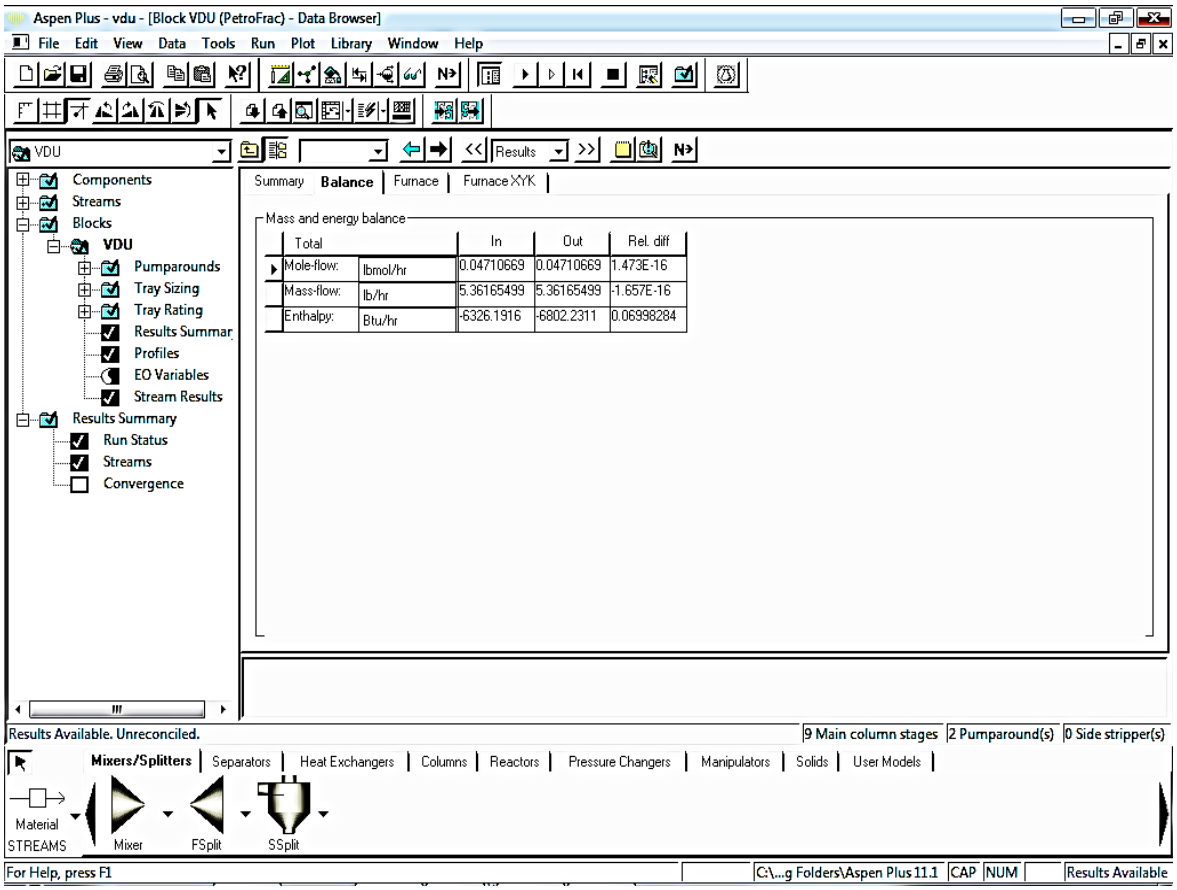


Figure F.21: VDU mass and energy balanc

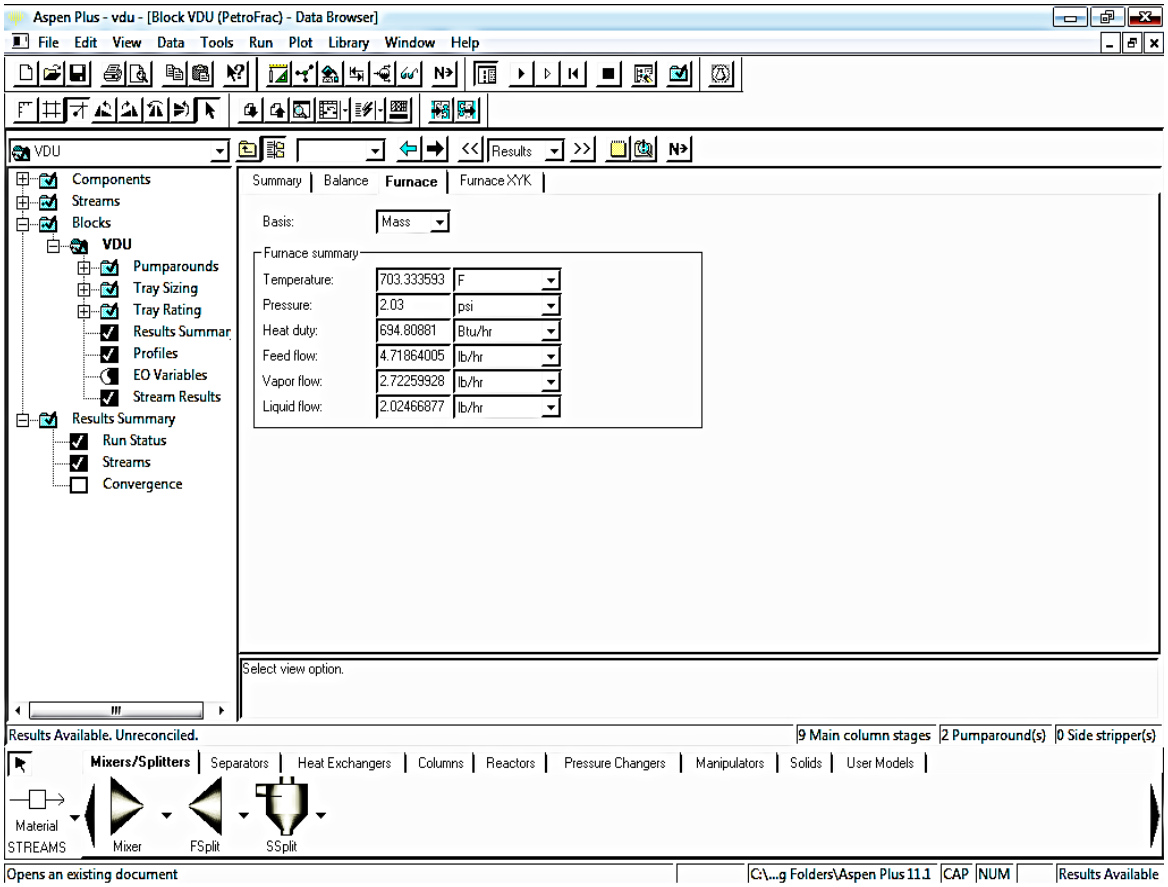


Figure F.22: VDU Furnace result

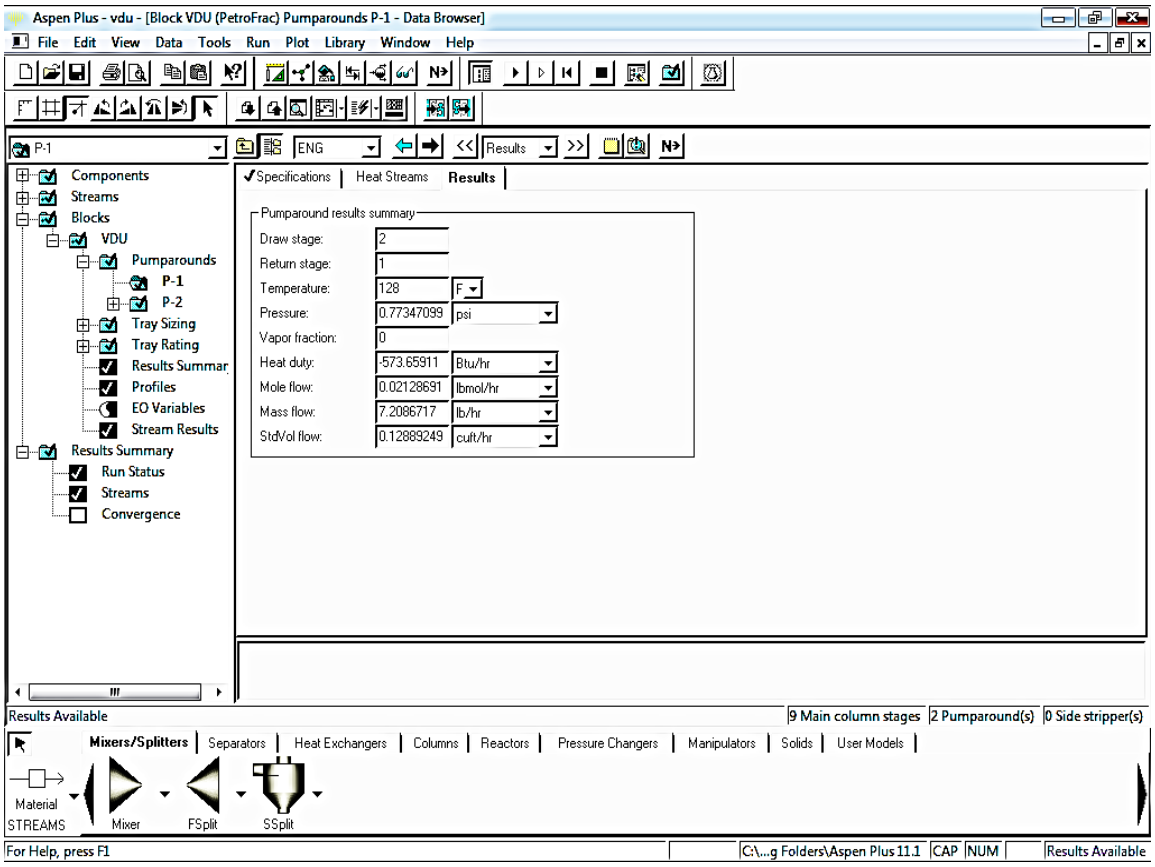


Figure F.23: VDU Top Pump Around result

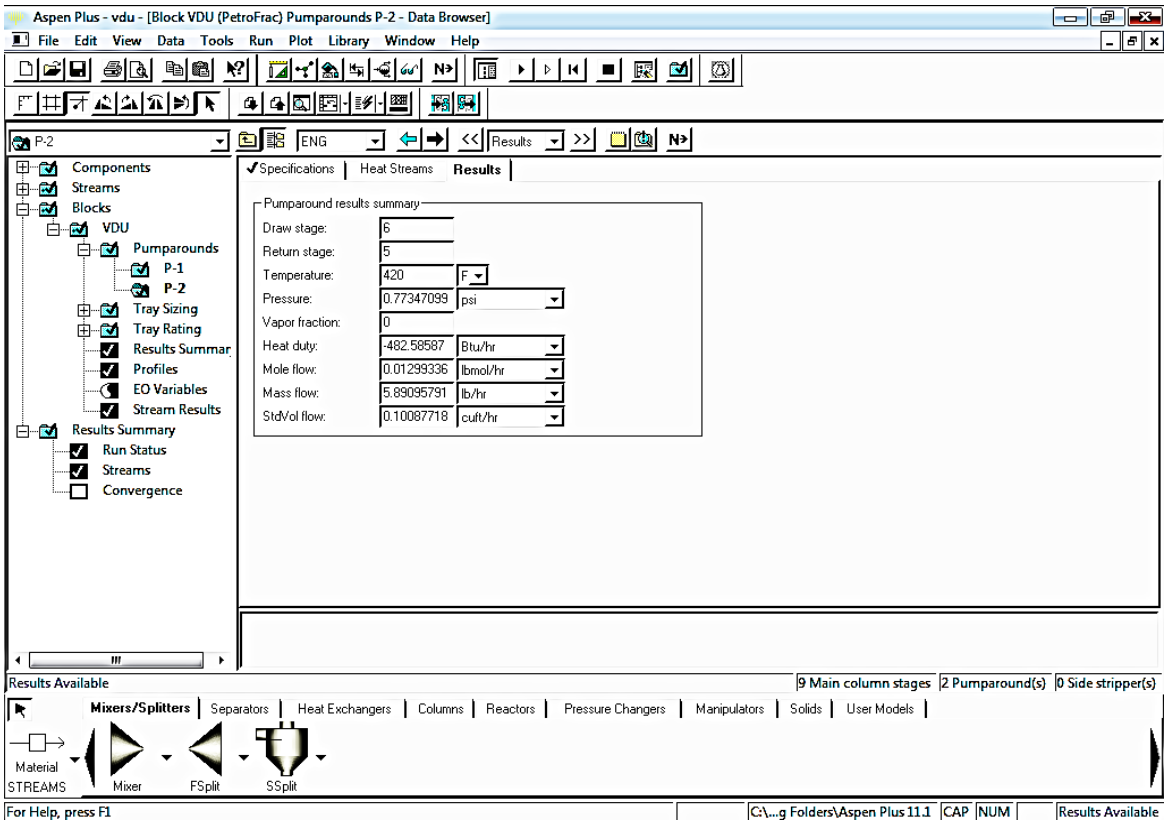


Figure F.24: VDU Bottom Pump Around results

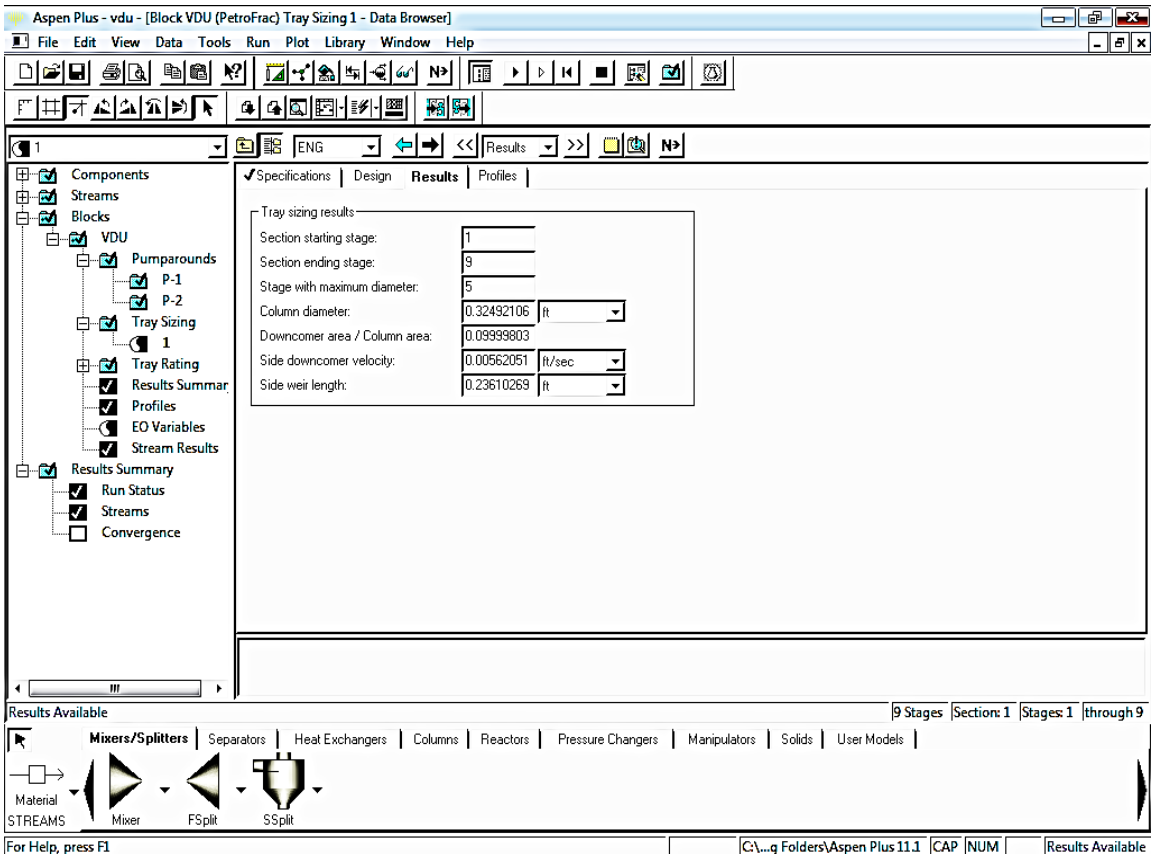


Figure F.25: Tray sizing result

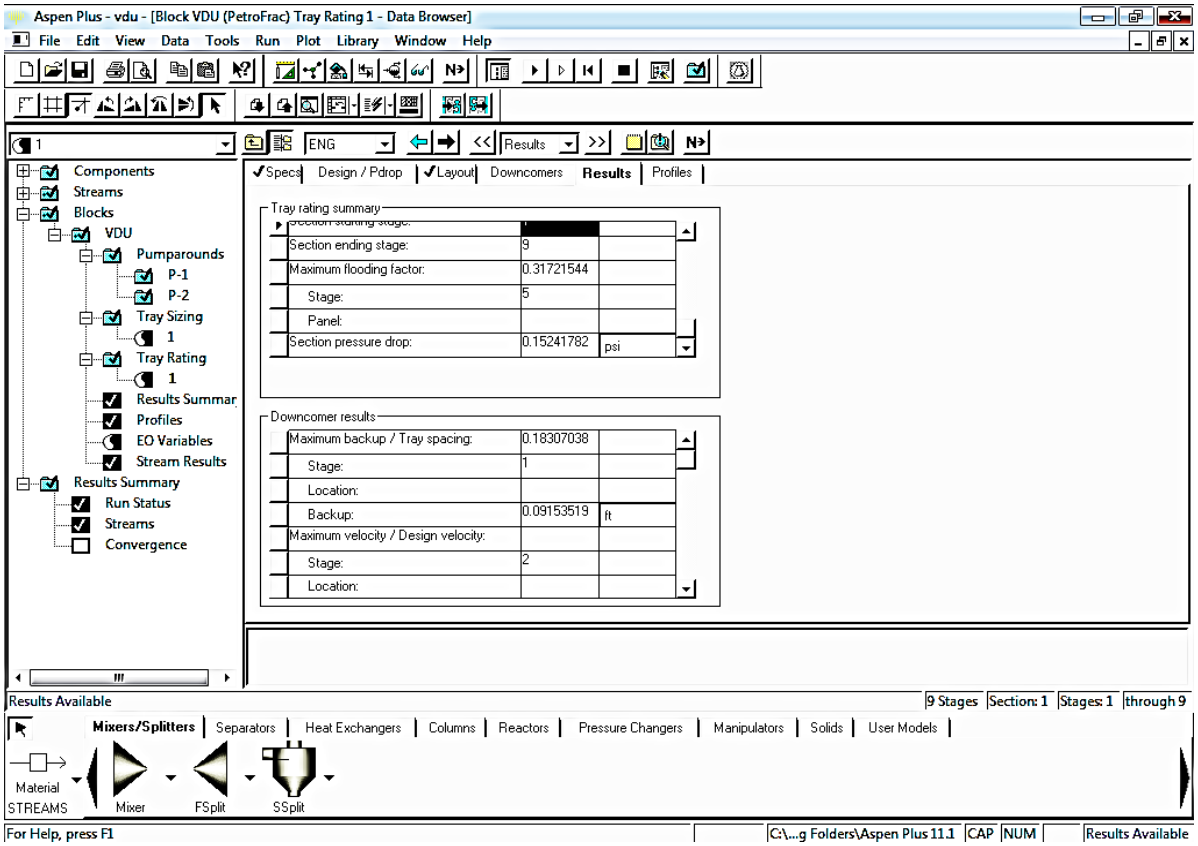


Figure F.26: VDU tray rating result

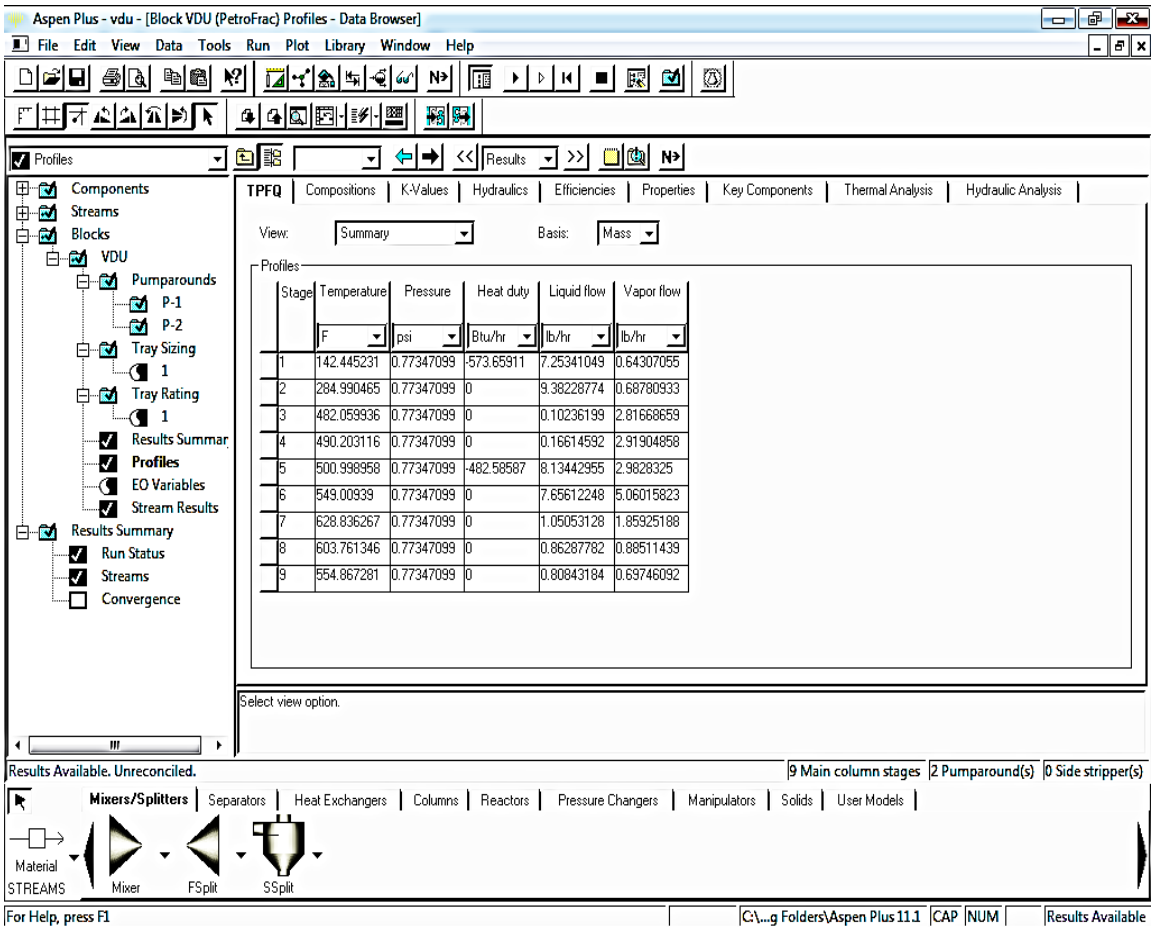


Figure F.27: VDU Profile result.

GLOSSARY

Absolute Pressure: (1) The reading of gage pressure plus the atmospheric pressure. (2) Gage pressure plus barometric or atmospheric pressure. Absolute pressure can be zero only in a perfect vacuum.

API; American Petroleum Institute

API gravity; The standard adopted by API for measuring the density of a liquid, (especially hydrocarbons) expressed in degrees. It can be converted from specific gravity by the following equation: Degrees API gravity = (141.5/specific gravity @ 60°F) -131.5

ASTM; American Society for Testing and Materials

Atmospheric Distillation: A basic refining process in which crude oil is heated in order to break it down into a number of intermediate components from which finished products can be made.

bar; Unit of pressure

Barrel; A quantity of 42 US Gallons (34.97UK Gallons).The traditional unit of measure of oil volume. $1m^3$ oil = 6.29 Barrels of oil

bbl; barrel(s)

bbl/d; barrel(s) of oil per day

bpd; Barrels per day.

Btu; British thermal unit

Boiling Range: The spread of temperatures, usually expressed in degrees Fahrenheit, over which oil starts to boil or distill vapors and proceeds to complete evaporation. Boiling range is determined by ASTM test procedures for specific petroleum products.

Boiling Temperature: Temperature at which steam bubbles begin to appear within a liquid. When the fluid is a pure compound, the boiling point is unique for each pressure.

°C; Degrees Celsius-°C = 5/9(°F-32)

Calorific value; The quantity of heat produced by complete combustion of unit weight of a material. Expressed as either calories per gram, or British Thermal Units (btu) per pound, or btu per standard cubic foot of gas.

Centipoise: (1) Viscosity measurement, 1/100th of a poise. (2) A centipoises (cP) is 1/100th of a poise (P), which is the fundamental unit of dynamic viscosity in the CGS system of units. In the SI unit of systems, the fundamental unit of dynamic viscosity is the pascal second (Pa s), where 1 Pa s is equivalent to 10 P.

Condenser: Heat exchanger in which vapor is liquefied by the rejection of heat to a heat sink.

Convection: Transfer of heat by a fluid moving by natural variations in density.

Corrosion: (1) The deteriorating chemical reaction of a metal with the fluids with which it is in contact. (2) The gradual decomposition or destruction of a material by chemical action, often due to an electrochemical reaction. Corrosion may be caused by (1) stray current electrolysis, (2) galvanic corrosion caused by dissimilar metals, or (3) differential-concentration cells. Corrosion starts at the surface of a material and moves inward.

Crude Oil; An unrefined mixture of naturally-occurring hydrocarbons. Because it is essentially a mixture, the density and properties of Crude Oil vary widely. Light Crude normally has an A.P.I. gravity of 30°or more. Gravities of 20° to 30° include the medium gravity crudes, while those below 20° are known as Heavy. Heavy oils are found right down to the residual solid state. Sour crude has significant sulphur content; Low sulphur crude is described as sweet.

Distillates; The products of distillation.

Distillation; The process of heating and "flashing" or boiling off successive fractions (component hydrocarbon substances) from a crude oil feed stock, or a product of earlier distillation.

Emulsion; A mixture in which one liquid, termed the dispersed phase, is uniformly distributed (usually as minute globules) in another liquid, called the continuous phase or dispersion medium. In an oil-water emulsion, the oil is the Dispersed phase and the water the dispersion medium; in a water-oil emulsion, the reverse holds. Emulsion is a typical product of oil wells

°**F**; Degrees Fahrenheit.1°F= 0.556 °C

Feedstock: Crude oil, natural gas liquids, natural gas, or other materials used as raw ingredients for making gasoline, other refined products, or chemicals.

Flaring/Venting: The controlled burning (flare) or release (vent) of natural gas that can't be processed for sale or use because of technical or economic reasons.

Flashing: Vaporization of water or light ends as pressure is released during production or processing.

Flow: Very simply, a movement of a fluid.

Fluid; A generic term meaning gas, vapour, liquid, or a combination thereof.

Fractionating Column: A tall tower, fitted with perforated trays, in which fractional distillation of crude oil or its products is carried out.

Fractions: Refiner's term for the portions of oils containing a number of hydrocarbon compounds but within certain boiling ranges, separated from other portions in fractional distillation. They are distinguished from pure compounds that have specified boiling temperatures.

Fuel Oil; The "bottom" or "heavy end" of the barrel after removal of middle distillates and lighter fractions. There are various grades of Light and Heavy Fuel Oil.

Furnace: Part of a boiler or warm air heating system in which energy is converted to heat, as by burning fuel or by converting electrical energy.

Gas Oil: (1) A liquid petroleum distillate having a viscosity intermediate between that of kerosene and lubricating oil. It derives its name from having originally been used in the manufacture of illuminating gas. It is now used to produce distillate fuel oils and gasoline. (2) A fraction derived in refining petroleum with a boiling range between kerosene and lubricating oil. It derives its name from having originally been used in the manufacture of illuminating gas. Now supplies distillate-type fuel oils and diesel fuel, also cracked to produce gasoline.

Gasoline: (1) The light fuel used to spark-ignition engines in cars, motorcycles, etc. Modern gasolines are blends of petroleum liquids that are produced in several different processes and that generally contain additives. (2) A refined petroleum naphtha that by its composition is suitable for fuel in a reciprocating-type internal combustion engine. ASTM D439 specifies three grades for various types of motor vehicle operations. Straight-run gasoline is the product of distillation; cracked gasoline that of a cracking process. (3) A light hydrocarbon mixture having C4–C9 carbon atom hydrocarbons, which are used as fuel for internal combustion engines. (4) Normally C7– C10 fuel, with a flash point of −40.

GPH; Gallons per hour

Gravity Flow: Water or wastewater flowing from a higher elevation to a lower elevation due to the force of gravity. The water does not flow due to energy provided by a pump. Wherever possible, wastewater collection systems are designed to use the force of gravity to convey waste liquids and solids.

Heat Exchanger; A process vessel which typically uses the passage of one fluid through a set of internal tubes to heat up or cool down another fluid in which they are immersed. There are many different designs and uses.

Heat Transfer Coefficient: Coefficient describing the total resistance to heat loss from a producing pipe to its surroundings. Includes heat loss by conduction, convection, and radiation.

Heavy Ends; Heavy or residual fractions of a feedstock after distillation, etc. Sometimes referred to as the "bottom" or "heavy end" of the barrel.

Heavy Oil; See Crude Oil.

Hydrocarbons; Organic compounds of hydrogen and carbon, whose densities, boiling points and freezing points increase as their molecular weights increase. Although composed of only two elements, hydrocarbons exist in a variety of Compounds because of the strong affinity of carbon atoms for other atoms and for itself. The smallest molecules of hydrocarbons are gaseous; the largest are solids. Petroleum is a mixture of many different hydrocarbons.

Ib; pound; the plural is also lb (libra, libræ)

Kerosene: (1) A medium range (C9–C16), straight-chain blend of hydrocarbons. The flash point is about 60°C (140°F), the boiling point is 174°C–288°C, and the density is 0.747–0.775 g/cc. (2) A medium-light distillate from the oil refining process; used for lighting and heating and for the manufacture of fuel for jet and turboprop aircraft engines. (3) Any petroleum product with a boiling range between the approximate limits of 140°C and 270°C, which satisfies specific quality requirements. (4) A light petroleum distillate that is used in space heaters, cook stoves, and water heaters and is suitable for use as a light source when burned in wick-fed lamps.

Light Crude; Generally applied to crude oil with an API gravity of 30 degrees or over. See American Petroleum Institute.

Light Ends; The least dense, more volatile parts of a crude oil stream in distillation.

Litre; 1 litre= 1,000 cc(or cm3);1(UK)gallon = 4.546 litre

Naphtha; A collective name given to a range if distillate fractions covering heavy gasolines and some of the lighter kerosene distillates.

OD; Outside diameter

Paraffins; The alkane hydrocarbons.

Parts per Million: (1) The scale on which impurities and contaminants in oils, gases, and petrochemicals are measured. (2) Unit weight of the solute per million unit weights of the solution. A small correction factor is needed to convert to mg/liter. (3) A measurement of concentration on a weight or volume basis. This term is equivalent to milligrams per liter (mg/L), which is the preferred term.

Pascal: A unit of pressure equal to 1/100,000 of a bar.

Petroleum; Literally ‘rock oil’. Can include products, but normally means crude oil.

Pressure Relief Valve: A mechanical valve that opens at a preset pressure to relieve pressure in a vessel.

Processing Plant: A facility designed to separate substances or make new substances through chemical reactions, procedures, or physical actions.

Process Variable: A physical or chemical quantity that is usually measured and controlled in the operation of a wastewater treatment plant or an industrial plant.

Pump: A mechanical device used to create flow.

Refined Products: The various hydrocarbons obtained as a result of refining process separation from crude oil. Typical refined products are LPG, naphtha, gasoline, kerosene, jet fuel, home heating oil, diesel fuel, residual fuel oil, lubricants, and petroleum coke.

Refinery: (1) An installation that manufactures finished petroleum products from crude oil, unfinished oils, natural gas liquids, other hydrocarbons, and oxygenates. (2) A plant used to separate the various components present in crude oil and convert them into usable fuel products or feedstock for other processes. (3) A large plant composed of many different processing units that are used to convert crude oil into finished or refined products. These processes include heating, fractionating, reforming, cracking, and hydrotreating.

Refinery Gas: A non-condensate gas collected in petroleum refineries.

Residuum: Residue from crude oil after distilling off all but the heaviest components, with a boiling range greater than 1000°F.

Retention Time: (1) The time based on the separator volume and shape and the fluid flow rate that produced fluids spend in a separator. It is based on the time needed to separate an emulsion.

Reynolds Number: An experimental number used in fluid flow to predict the flow velocity at which the flow regime moves from laminar flow through a transition range and into turbulent flow. It is the dimensionless ratio of inertial forces to viscous forces in flowing fluids. It may be viewed as a ratio of the shear stress due to turbulence to the shear stress due to viscosity. Flow with a Reynolds number less than 2000–4000 is laminar flow; that with a Reynolds number greater than 2000–4000 is turbulent flow.

Separator; A cylindrical or spherical vessel used to separate the components in streams of mixed fluids.

Specific Gravity: (1) A measure of the density of a material usually obtained by comparing it with water. (2) The ratio of the density of a substance to the density of a reference substance, both at specified physical conditions. As applied to gas, air is the reference substance and the physical conditions are a specified temperature and atmospheric pressure. (3) The ratio of the density of a substance to the density of a comparison material, usually at a specific temperature and pressure. (4) The ratio of the weight of a given volume of a substance to the weight of an equal volume of water.

Stripping: In refining, the removal of the more volatile components from a cut or fraction in order to raise the flash point of kerosene, gas oil, or lubricating oil.

Sulfur: (1) A yellow mineral extracted from petroleum for making fertilizers, pharmaceuticals, and other products. (2) A nonmetallic element that occurs in association with salt diapirs throughout much of the onshore and offshore Gulf of Mexico region. All offshore sulfur is mined by the Frasch process, which uses hot brine to melt sulfur out of the enclosing rock so the molten sulfur can be recovered. (3) A yellowish white solid. Sulfur appears in oil and gas in the form of hydrogen sulfide or in combination with a hydrocarbon to form a mercaptan. Sulfur is an undesirable component because when the product is burned it forms sulfur oxides, which contribute to air pollution. (4) A yellowish nonmetallic element, sometimes known as "brimstone." It is present at various levels of concentration in many fossil fuels whose combustion releases sulfur compounds that are considered harmful to the environment. Some of the most commonly used fossil fuels are categorized according to their sulfur content, with lower sulfur fuels usually selling at a higher price.

Temp; Temperature

Throughput (Processing): Average amount of raw material that is processed in a given period by a facility, such as a natural gas processing plant, an oil refinery, or a petrochemical plant.

Tower: A refinery apparatus used in connection with a still to increase the degree of separation of fractions obtained during the distillation of oil in the still. Also called a column.

Vacuum Distillation: (1) Distillation under reduced pressure (less the atmospheric) that lowers the boiling temperature of the liquid being distilled. This technique with its relatively low temperatures prevents cracking or decomposition of the charge stock. (2) Distillation under reduced pressure, which reduces the boiling temperature of the material being distilled sufficiently to prevent decomposition or cracking.

Vapour pressure; The pressure exerted by the vapour from a substance.

Vent; A pipe or fitting on a vessel that can be opened to the atmosphere

Weir: A wall or plate placed in an open channel.

Yield; The total amount of product of a refinery process or of all products of all processes of a refinery compared with the equivalent amount of feedstock. A "Refinery Yield" for a given crude oil feedstock and refinery will include a table of amounts of all products derived from a barrel orton(ne) of crude minus the refinery's own usage for fuel, flaring and other losses. Since refineries commonly use several feedstocks simultaneously, such a yield is often measured or predicted on the basis of adding incremental feedstock to a fixed throughput.