Sudan University of Science and Technology

College of Graduate Studies

Upgrading of Heavy Crudes and Residue for Liquid fuels production

Case Study: EL Obeid Refinery Company

تحسين الخامات الثقيلة والمتبقي لإنتاج الوقود السائل

دراسة حالة: شركة مصفاة الابيض

A Thesis Submitted in Fulfillment for the Requirements of the Degree of Doctor of Philosophy in chemical Engineering

By

YASIR ABDALLA DAFAALLA ALI

Main Supervisor
Professor. Gurashi Abdalla Gasmelseed
Co Supervisor
Dr. Tomadir Awad Ibraheem Hamed

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آية قرآنية

{ وَمَا تَوْفِيقِي إِلَّا بِاللَّهِ عَلَيْهِ تَوَكَّلْتُ وَإِلَيْهِ أُنِيبُ }

صدق الله العظيم

DEDICATION

I dedicate this work to the great late professor Hamid Mohamed Mustafa Ahmed, that wonderful person from whom I learned a lot. I ask God to grant him Paradise;

To my parents, whose encouragement was a great source of inspiration to me;

To my wife, for her patience and steadfast support

And lastly but not least to my lovely kids, Gofran and Karar in whose intelligent and bright eyes I see my own unfulfilled dreams coming true in the neither too far nor too distant future.

.

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Last but not least, I want to express my great gratitude to my parents for their enormous love, my brothers and my sisters for their unconditional support and encouragement.

My warmest thanks go to my wife, my son and my daughter for their love, understanding, and patience, during my study.

ABBREVIATIONS

API	American Petroleum Institute
ENI	Ente Nazionale Idrocarbure
HDT	Hydrotreating
NEBULA^{TM}	New Bulk Activity, Exxon Mobil
STARS^{TM}	Super Type II Active Reaction Sites
EMRE	Exxon Mobil Research &Engineering
SYDEC	Selective Yield Delayed Coking
CLG	Chevron Lummus Global
OCR	On-stream Catalyst Replacement technology
UFR	Up Flow Reactor
UNITAR	United Nation Institute for Training And Research
TBP	True Boiling Point
SBD	Spinning Band Distillation
ASTM	the American Society for Testing and Materials
ADC	Advance Distillation Curve
DVLA	Deep Vacuum Laboratory Apparatus
PDF	Probability Distribution Function
PIONA	Paraffin, Iso-paraffin, Olefin, Naphthene and Aromatic
SARA	Saturates Aromatics Resins and Asphaltenes
TGP	Texaco Gasification Process
THGP	Texaco Hydrogen Generation Process
TGPS	Texaco Gasification Power Systems
DAO	De Asphalted Oil
VDU	Vacuum Distillation Unit
ORC	Obeid Refinery Company
CDU	Crude Distillation Unit
AGO	Atmospheric Gas Oil
LVGO	Light Vacuum Gas Oil
HVGO	Heavy Vacuum Gas Oil
LPG	Liquefied Petroleum Gas
P	Proportional
PI	Proportional Integral
PID	Proportional Integral Derivative

ABSTRACT

Trends have predicted that peak oil production, will be reached around 2020-2030, where the production of oil starts to decline after that, there will be less and less oil to refine, the depletion of light oil and the severe shortage of light products is what compelled us to search for solutions to provide the necessities, observe that the crude that feeds the refineries have become more heavy, so it was necessary to develop oil refining processes to meet the production requirements in quantity and quality.

This study aims to search for solutions to the problem of heavy sudanese crude oil and residues to produce light products to fill the shortage, and El Obeid refinery was taken as a case. A detailed simulation of an atmospheric distillation column was performed using Aspen Hysys v.10 software. The model was based on data from the actual plant located in El Obeid refinery in North of Kordofan. Crude oils were described using True Boiling Point (TBP) assays and a Peng-Robinson package was used to predict thermodynamic properties. Three different types of crude oils (Nile Blend, Rawat, Thargath), and 7 blends with different proportions were selected from the previous crude oils.

The simulation results agree well with those of the industrial plant in the case of the Nile blend and the validity of the model is confirmed. The conversion ratio to high-value light products in the atmospheric distillation tower of the Nile blend ware 31%vol, Rawat were 39.7%vol, Thargath were 24.22%vol, Mix 1 were 30.51%vol, Mix 2 were 26.62%vol, Mix3were 31.23%vol, Mix 4 were 28.96%vol, Mix 5 were 30.47%vol, Mix 6 were 27.97%vol and Mix 7 were 27.81%vol.

It was proposed to add a new unit (atmospheric distillation tower) identical to the first distillation tower to increase the inputs to the second proposal (vacuum distillation tower) to convert the residues of towers into light products, four streams were selected from to feed their residues to the vacuum distillation tower,

and the residues were converted to high-value light products. The selected streams were (Nile Blend, mix 1, mix 3 and mix 7) and the percentage of convertion into light products were (52.57%, 57.48% ,62.57% and 66.94%) by volume respectively.

These results clearly indicate that blending the Sudanese heavy crudes improves their properties and gives better desired products than Nile blend as in Mix 1, Mix 3 and Mix 7.

ملخص الأطروحة

تنبأت بعض النظريات بأن ذروة إنتاج النفط ستكون بين 2020-2030 م، حيث يبدأ إنتاج النفط في الإنخفاض مما يقلل من كمية النفط المستخدم في التكرير، نضوب النفط الخفيف والنقص الحاد في المنتجات النفطية الخفيفة هو ما دفعنا للبحث عن حلول لتوفير الضروريات، من الملاحظ أن الخامات أصبحت ثقيلة جداً، لذلك أصبح من الضروري تطوير عمليات تكرير النفط لتلبية متطلبات الإنتاج كماً ونوعاً.

هدفت هذه الدراسة إلى البحث عن حلول لمشكلة النفط الخام السوداني الثقيل ومخلفاته وإنتاج منتجات خفيفة لسد النقص، تم إجراء محاكاة مفصلة لعمود التقطير الجوي بإستخدام برنامج آسبن هايسس الإصدار 10، إعتمد النموذج على بيانات من المصفاة الفعلية الواقعة في مدينة الأبيض بشمال كردفان.

تم وصف الزيوت الخام بإستخدام معادلات الحالة للتنبؤ بالخصائص الديناميكية الحرارية، تم إختيار ثلاثة أنواع مختلفة من الزيوت الخام (مزيج النيل، خام الراوات وخام ثارجاث)، ومن ثم تم مزج الخامات الثلاث بنسب مختلفة لتكوين خامات أخرى وتم تكوين سبعة خلطات.

تتفق نتائج المحاكاة بشكل جيد مع نتائج المصفاة في حالة مزيج النيل وتأكد صحة النموذج. كانت نسبة التحويل إلى المنتجات الخفيفة عالية القيمة في برج التقطير الجوي لمزيج النيل 31٪ حجمًا، الراوات39.7 ٪ حجمًا، ثارجاث24.22 ٪حجمًا، الخليط 1 كان 30.51٪ حجمًا، الخليط 2 كان 26.62٪ حجمًا، الخليط 3 كان 31.23٪ حجمًا، الخليط 4 كان 28.96٪ حجمًا، الخليط 5 كان 30.47٪ حجمًا، ا الخليط 6 كان 27.97٪ حجمًا والخليط 7 كان 27.81٪ حجمًا.

تم اقتراح إضافة وحدة جديدة (برج التقطير الجوي) مطابقة لبرج التقطير الأول لزيادة المدخلات للاقتراح الثاني (برج التقطير الفراغي) لتحويل متبقي الأبراج إلى منتجات خفيفة، تم اختيار أربعة خامات منها لتغذية بقاياها إلى برج التقطير الفراغي، وتم تحويل البقايا إلى منتجات خفيفة عالية القيمة. الخامات المختارة كانت (مزيج النيل، خليط 1، خليط 3، خليط 7) ونسبة تحويلها إلى منتجات خفيفة (52.57٪، 57.48٪، 62.57٪، 66.94٪) بالحجم على التوالي.

تشير هذه النتائج بوضوح إلى أن مزج الخامات السودانية الثقيلة يحسن خواصها ويعطي منتجات مرغوبة أفضل من مزيج النيل كما هو الحال في خليط 1، خليط 3، خليط 7.

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CHAPTER ONE

Introduction

1.1 Background

Light crude oil is liquid petroleum that has low density and that flows freely at room temperature. It has low viscosity, low specific gravity and high API gravity due to the presence of a high proportion of light hydrocarbon fractions. It generally has a low wax content as well. On the other hand, heavy crude oil or extra heavy crude oil is oil that is highly viscous and cannot easily flow under normal conditions. Heavy crude oil has been defined as any liquid petroleum with American Petroleum Institute (API) gravity less than 20°API. Extra heavy oil is defined with API gravity below 10°API. Physical properties that differ between heavy crude oils and lighter grades include higher viscosity and specific gravity, as well as heavier molecular composition.

Light crude oil receives a higher price than heavy crude oil on commodity markets because it produces a higher percentage of gasoline and diesel fuel when converted into product by an oil refinery. Heavy crude oil has more negative impact on the environmental than its light counterpart since its refinement requires the use of more advanced techniques and the use of contaminated.

As world demand for petroleum products increases, and production in existing light crude fields declines. heavier, high sulfur crude oils are increasing in market share. Since heavy sour crudes typically trade at 5-15$ per barrel discount to light sweet crudes. It may be profitable for existing refinery to increase their throughput of heavy sour crudes.

The percentage of heavy crudes produced has been increasing in recent years as per data published by the (Ente Nazionale Idrocarbure) ENI Group

(World Oil and Gas, 2006) and the trend is for decreasing supplies of light and medium crude and increasing supplies of heavy crude.

The heavier crude oils are having a higher sulfur content thus the quality of crude oil has been worsening, in term of sulfur content. Practically all of the heavy crude oil produced in recent years was either medium sour (0.5-1.0% sulfur) or completely sour (more than 1% sulfur).

It is also noticed that while globally, sweet crudes account for about 30% of present production; they comprise only about 20% of world oil reserves. The middle East- which is the largest potential source of crude oil produces almost no sweet crude oil-about $500*10^3$ barrel/day out of $20*10^6$ barrel/day total crude oil production.

Crude oils are getting heavier or even extra-heavy with higher sulfur, nitrogen and metal contents (F.E. Biasca *et al*.,2003); in the meantime, the demand for lighter products particularly gasoline ends (<150 °C) is growing. It became necessary to convert bottom of the barrel into cleaner and more valuable liquid products. The bottom of the barrel residue represents a significant portion (up to 85vol %) of a barrel of crude.

The environmental concerns have increased, resulting in more rigorous specifications for petroleum products. These stringent and rigid environmental regulations forced the refiners to seek ways and technology for production of lighter products of higher quality and fewer contaminants. Thus most of the technological development is presently directed towards improving refinery upgrading processes (Alan G. Lucas, 2001).

The demand for high value petroleum products such as middle distillate, gasoline and lube oil is increasing, while the demand for low value products such as fuel oil and residue based products is decreasing.

Therefore, maximizing of liquid products yield from various processes and residues is of immediate concern to refiners(Hamed, 2009).

In the last century, catalysis became one of the most powerful tools in the petroleum sectors. Although it is often mentioned that the field of hydroprocessing catalysis is mature and there are not much compasses for researcher, the increasing demand of heavy oil has become hydroprocessing.

A challenging task for refiners as well as for researchers. During the last 50 years, Hydrotreating (HDT) catalysis has been recognized as a multidisciplinary work about the support effect, catalyst characterization, and catalytic activity for deep hydrodesulphurization (Topsøe H *et al.*, 1996).

Now, when refiners are already using highly active catalyst like NEW BULK Activity, ExxonMobil (NEBULA^{TM}) or Super Type II Active Reaction Sites, AKZONoble Catalyst (STARS^{TM}) for middle distillates (close to fulfill the EPA sulfur specifications) (Pappal DA *et al*., 2003), the researcher's importance is increasing towards the heavy oil hydroprocessing.

The HDT catalyst for heavy oil is entirely different than those for model molecules and middles distillate. Therefore, the importance of the catalyst is not only focused on its development, but also tremendous work has to be done for catalyst formulation as well as process designing.

A number of technologies have been developed over the years for heavy crude and residue oil upgrading, which include processes that are based on carbon rejection, hydrogen addition and combination of both routes, (http://www.spe.org/spesite/spe/spe/industry/reserves/petroleumReserves Definitions 1997);(http://www.researchandmarkets.com/research/5787de/heavy crude oil: Outlook to 2020) most of them with proven commercial experience. Those technologies are illustrated below:

1.2 Carbon Rejection

Carbon rejection is one of the first types of conversion processes applied in the oil industry, and has been used since1913 for different fuels and heavy hydrocarbons heated under-pressure. Carbon rejection processes represent

56.6% of the total worldwide processing capacity mainly due to its relative low investment Carbon rejection technologies include processes such as visbreaking, thermal cracking, and coking. Foster Wheeler and Universal Oil Products made an agreement to combine their technical experiences, developing over 50 visco-reduction plants. Shell and ABB-Lummus have developed and commercialized a drum type application namely Reaction Chamber (Soaker), with over 80 projects based on this process (Refining Processes 2000) ;(http://www.cbi.com/images/uploads/tech sheets/Visbreaking.pdf, Shell Soaker Visbreaking ,2009). Fluid coking and Flexicoking are developments of Exxon Mobil Research & Engineering (EMRE).

The combined capacity of Fluid coking and Flexicoking amounts241*10^3 and 426*10^3bbl/day, respectively. Delayed Coking a mature technology is offered by Foster wheeler Selective Yield Delayed Coking (SYDEC) with over 25 revamps designed in the last 30 years, and over 20 new units designed in the last 20 years (SYDEC delayed coking.2009).

1.3 Hydrogen Addition

Traditional upgraders reduce carbon-to-hydrogen ratio by adding hydrogen obtained from natural gas. Hydrogen addition technologies are classified depending on the type of reactor used, e.g. fixed-bed, moving-bed, ebullated-bed and slurry-bed processes. Axens technology utilizing its Hyvahl process (Permutable Reactor System), has reached an important commercialization level. Chevron Lummus Global (CLG) licenses On-stream Catalyst Replacement technology (OCR) for processing high-metal feeds and the revamp alternative, an Up Flow Reactor (UFR).

There are also other technologies at different level of development or close to commercialization or even already used, that have been abandoned as uncompetitive during the years of cheap crude oil. The changing prices of crude oils as well as the increasing production of heavy and extra-heavy crude oils have motivated more research and development aiming at upgrading of such heavy materials.

1.4 Problem Statement

As the world supply of light sweet crude oil is depleted, the stocks of heavy oil and bitumen become most important as a component in supplying the demand for fuels and petrochemical feed. Similar the need to process the residue fractions of petroleum has also increased in importance.

Thirty years ago, the process option was limited; residue was used for coking, manufacture of asphalt, or refinery fuel. the combination of higher fuel prices and environmental constraints has given rise to a new generation of process alternative and focused attention on the efficient upgrading of residue, in order to meet the challenge of converting the high molecular weight component of residue and heavy oil in to distill-able fraction, one requires acknowledgment of chemistry of these feed, their physical and thermodynamic properties, and the available conversion process, Today, heavy oils, extra-heavy oils and bitumen represent only a small portion of worldwide oil production, and available solution to use these standard resources ,Can be represented by development of methods of extraction the light product from the heavy crudes and residue.

The proportion of heavy crude oil increased in the recent period and demand is on the light products, and it became necessary to upgrading heavy crude and residues to produce the desired product.

1.5 Objective of Study

- Convert marginal heavy crude oil into light, higher value crude, and convert heavy, sour refinery bottoms into desired light product.
- Investigation of the different heavy crude and residue processing technique to select the most appropriate and determine the suitability of such method to local refineries.
- using the specialized applications in simulation, design and control to Improving refinery upgrading processes.

CHAPTER TWO

Literature Review

An understanding of the nature of crude oil, oil refinery processes, oil characterization, asphaltene precipitation, and phase behavior modeling is required in order to be able to model asphaltene precipitation in downstream processes. Each of these topics is reviewed below.

2.1 Crude Oil Definition

Hydrocarbons occur naturally as gas, gas condensate, crude oil from light oils to extra-heavy oils, bitumen, and coal. Crude oils are classified in terms of their physical properties (viscosity and density/API) as indicated with the United Nation Institute for Training And Research (UNITAR) classification given in Table 2.1.

Table 2.1. UNITAR classification of oils by their physical properties at 15.6^oC *(Gray, 1994)***.**

Property **Type**	**Viscosity mPa.s**	**Density** $\mathbf{kg/m^3}$	**API Gravity**
Conventional oil	<102	< 934	>20^o
Heavy oil	102 – 104	934 – 1000	20^o – 10^o
Bitumen	>104	>1000	<10^o

Crude oil is a mixture of hundreds of thousands of different hydrocarbon based molecules many of which also contain non-hydrocarbon components such as nitrogen, oxygen, sulfur and sometimes organometallic compounds. Crude oil can be thought of as a distribution of paraffinic, naphthenic, and aromatic species of increasing molecular- weight and complexity as described in (Altgelt *et al* .1994).

As seen in Figure 2.1, crude oil boiling point increases as the complexity of the components making up the oil increases. The left boundary (Paraffin's solid line) in Figure2.1 represents the boiling point of paraffin's (*n*-alkanes) for which molar mass rapidly increases at higher boiling points. The right boundary at higher boiling points (>300°C for Polar Polyfunctional compounds dash line) in Figure 2.1 represent complex multi-aromatic ring structures and Polyfunctional compounds. The largest and most complex components are the Asphaltenes which make up 5 to 20 wt% of native heavy oils and bitumen and up to 50wt% or 60wt% of vacuum residues in refineries.

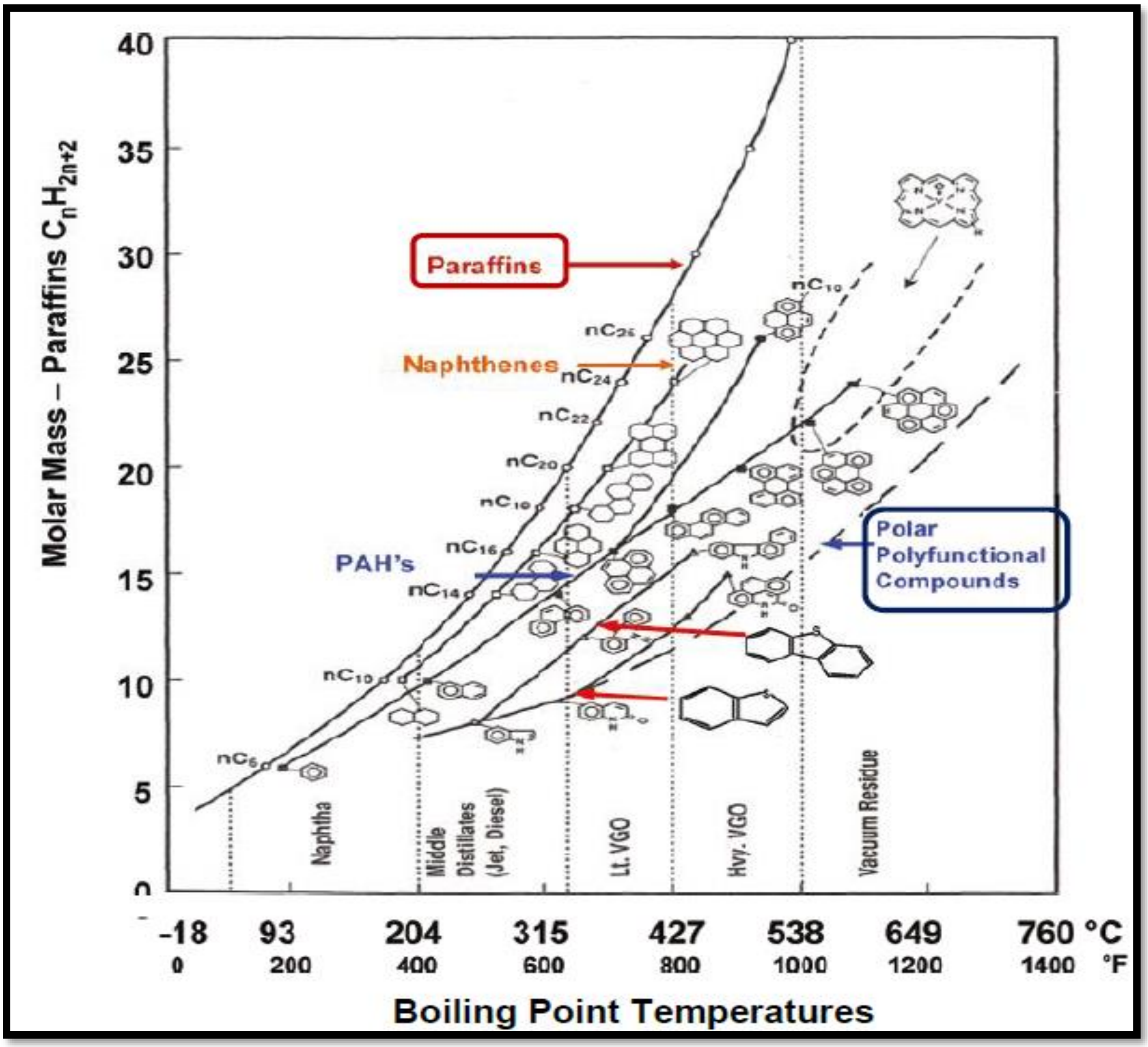


Figure 2.1: Effect of molecular weight and structure on boiling (Altgelt and boduszynsi,1994)

In comparison with conventional oils, heavy oils have high viscosity, high asphaltene content, and relatively high heavy metals, sulfur, and nitrogen contents, usually out of market specifications. Heavy oils are further classified into three different categories according to specific gravity and viscosity at reservoir conditions (Huc, 2011):

- Heavy oil: API degree higher than 10, ambient viscosity lower than 10,000 cP, and flows at reservoir conditions
- Extra-heavy oil: API degree less than 10, in situ viscosity less than 10,000 cP, some mobility at reservoir conditions
- Natural Bitumen: API degree less than 10, in situ viscosity greater than 10,000 cP, and does not flow at reservoir conditions.

Conventional oils, bitumen, and vacuum residues are used in this project as well as in situ reacted oils and converted oils from a visbreaker and a hydrocracking process.

2.2 Desalting Crude Oils

If the salt content of the crude oil is greater than 10 lb/1000 bbl (expressed as NaCl), the crude requires desalting to minimize fouling and corrosion caused by salt deposition on heat transfer surfaces and acids formed by decomposition of the chloride salts. In addition, some metals in inorganic compounds dissolved in water emulsified with the crude oil, which can cause catalyst deactivation in catalytic processing units, are partially rejected in the desalting process.

The trend toward running heavier crude oils has increased the importance of efficient desalting of crudes. Until recently, the criterion for desalting crude oils was 10 lb salt/1000 bbl (expressed as NaCl) or more, but now many companies desalt all crude oils.

Reduced equipment fouling and corrosion and longer catalyst life provide justification for this additional treatment. Two-stage desalting is used if the crude oil salt content is more than 20 lb/1000 bbl and, in the cases where residua are catalytically processed, there are some crudes for which three-stage desalting is used.

The salt in the crude is in the form of dissolved or suspended salt crystals in water emulsified with the crude oil. The basic principle is to wash the salt from the crude oil with water. Problems occur in obtaining efficient and economical water/oil mixing, water-wetting of suspended solids, and separation of the wash water from the oil. The pH, gravity, and viscosity of the crude oil, as well as the volume of wash water used per volume of crude, affect the separation ease and efficiency.

A secondary but important function of the desalting process is the removal of suspended solids from the crude oil. These are usually very fine sand, clay, and soil particles; iron oxide and iron sulfide particles from pipelines, tanks, or tankers; and other contaminants picked up in transit or production. Total suspended solids removal should be 60% or better with 80% removal of particles greater than 0.8 micron in size (A. A. Kutler, 1969).

Desalting is carried out by mixing the crude oil with from 3 to 10vol% water at temperatures from 90 to 150°C. Both the ratio of the water to oil and the temperature of operation are functions of the density of the oil.

Table2.2: Explain to Typical operating conditions (Gary, J.H and G.E. Handwerk,2001)

API	Water wash Vol %	Temperature F° (C °)
>40	3-4	240-260 (115-125)
30-40	4-7	260-280 (125-140)
<30	7-10	280-330 (140-150)

The salts are dissolved in the wash water and the oil and water phases separated in a settling vessel either by adding chemicals to assist in breaking the emulsion or by developing a high-potential electrical field across the settling vessel to coalesce the droplets of salty water more rapidly. Either AC or DC fields may be used and potentials from 12,000 to 35,000 volts are used to promote coalescence. For single-stage desalting units 90 to 95% efficiencies are obtained and two-stage processes achieve 99% or better efficiency.

One process uses both AC and DC fields to provide high dewatering efficiency. An AC field is applied near the oil–water interface and a DC field in the oil phase above the interface. Efficiencies of up to 99% water removal in a single stage are claimed for the dual field process. About 90% of desalters use AC field separation only.

2.3 Crude Oil Conversion Processes

Crude oils, heavy oils, and bitumen are refined to obtain high-value products, such as liquefied fuels, solvents, lubricants and petrochemical streams (Speight, 1998). The basic refinery process is distillation which fractionates the crude oil into products of differing volatility. After distillation some streams are further fractionated by other physical methods such as absorption, adsorption, and solvent extraction. The bottom products from these separations (usually atmospheric or vacuum residua) are sent to conversion processes to increase the yield of lighter higher value products. These conversion processes are based on either carbon rejection or hydrogen addition reactions.

Carbon rejection processes, such as visbreaking, involve the redistribution of hydrogen among the various components via thermal or catalytic cracking. During carbon rejection processes, the concentration of high molecular weight and high polarity material increases until a sufficient concentration is reached to form a separate dense, aromatic liquid phase termed coke (Trejo *et al.*, 2009). Hydrogen addition processes, such as hydroconversion (hydrocracking or hydrotreatment), supply an external source of hydrogen to increase the hydrogen to carbon atomic ratio (H/C) of the feedstock (Ancheyta *et al.,* 2010). Hydrogen addition processes decrease the yield of coke formation (Rana *et al.,*

2007) and increase the H/C ratio in comparison with carbon rejection processes. For both carbon rejection and hydrogen addition processes, the severity of the treatment (reaction time and temperature) determines the extent of reaction and the final product composition.

During the last few years, many approaches to upgrading bitumen and heavy oils and improving upgrading process efficiency have been developed and some of them have been implemented in the oil industry. A brief description of the main processes for heavy oils and residues is provided below.

2.3.1 In-Situ Upgrading

In situ upgrading, usually referred to as in-situ heat treatment, is a process to reduce the viscosity and density of the oil in a subsurface formation. In-situ upgrading consists of thermal treatment of the oil in the reservoir using a heating device such as electrical heaters. During in-situ heat treatment, pyrolysis reactions occur where high temperature is maintained in the formation. This process produces a lower density, lower viscosity oil, synthesis gas (H_2, CO_2, H_2O, N_2, CH_4 and other gases in lower proportions), and a coke-like residue. The amount and properties of the products depends on the temperature and the type of oil (Roes *et al.,* 2009).

In situ combustion processes can also upgrade part of the in-situ oil.

In-situ combustion consists of injecting an oxidizing agent, such as air or oxygen enriched air, into the reservoir to burn a portion of the oil and promote mobility and higher production of the remaining oil. Part of the oil close to the burning zone is also cracked into lighter components. The oxidizing agent is injected for “short” periods of time (days or few weeks) sometime with the aid of steam (Castanier and Brigham, 2003). Mass transport and several chemical reactions occur as the process progresses and temperature gradients develop in the reservoir. Some of the chemical reactions are combustion, oxidation, cracking, and pyrolysis (Castanier et al., 2003; Jia et al, 2013).

In-situ combustion can also lower sulfur, metals and asphaltene content depending on the conditions and additives added during the operation.

2.3.2 Thermocracking

The main objective of thermocracking processes is to mildly crack the heavy hydrocarbon molecules (high molecular weight) to produce more distillates (naphtha and gasoline) from atmospheric and vacuum residues, extra heavy oils, and bitumen. The main processes are visbreaking, catalytic cracking, and coking.

Visbreaking, one of the basic processes in refining resids, is a non-catalytic thermal process used to reduce the viscosity of residue fractions and produce limited amounts of distillables. It is an inexpensive upgrading process especially used for residues of conventional oils and bitumen (Lababidi *et al*., 2014; Xu *et al*., 2013). The operating conditions of a visbreaker are critical to attain a high degree of conversion without worsening the stability of the resid. Operating conditions are also very important to attain large operation times before cleaning maintenance is required to remove coke deposits from the reactor (Vezirov, 2011).

Catalytic cracking uses catalyst to promote faster and more effective cracking reactions with higher cracking rates (Corma and Wojciechowski, 1985). Different type of catalyst beds have been developed according to the needs and research of major oil companies. The main reaction is thermocracking of carbon-carbon bonds into a lighter paraffin’s and olefin.

Coking is used to reduce the carbon-to-hydrogen ratio. Residues are transformed at higher temperatures than the visbreaking process. Coking is also used to reduce the viscosity of extra heavy oils in production fields so that the product can be easily transported to the refinery for further downstream treatment (Huc, 2011).

2.3.3 Catalytic Hydroconversion: hydrocracking & hydrotreatment

Catalytic hydrotreatment and hydrocracking involve thermal-processing of a feedstock, such as a vacuum residue, with a catalyst and hydrogen at pressure to produce lighter and more valuable products from a large variety of feedstocks (Altgelt *et al.*, 1994; Kaes, 2000; Speight, 1998). They are very important processes in the refinery because they convert heavy and/or sulfur and aromatic-rich feedstocks into middle distillates of high quality within market specifications including low sulfur, minimum impurities, and lower aromatic content (Minderhoud and van Veen 1993). The process is referred as hydrotreatment instead of hydrocracking when the main purpose is to reduce the organometallic compounds (hydrodemetalization) (Huc, 2011).

There are several configurations for commercial hydrocracking processes which can be classified into four main categories: fixed bed, moving bed (co-or counter-current), ebullated bed, and entrained bed reactors (Huc, 2011). One of the most common processes configurations is a two-stage process with one or more reactors for each stage. The first stage is usually where the cracking reactions occur; the second stage serves to reduce sulfur, nitrogen, oxygen and aromatic content of the feedstock. Hydrocracking processes play an important role converting bitumen, heavy feedstock’s or resids into lighter and more valuable synthetic oils and are nowadays used by major oil companies in Alberta and worldwide (Huc, 2011; Roes *et al.*, 2009).

2.4 Crude Oil Characterization

Crude oil characterization is the identification and quantification of the main compounds or pseudo-components making up the crude oil mixture. Characterization of crude oil feedstocks and product streams is required for the design and operation of all refinery processes including conversion processes. One objective of the characterization is to:

- evaluate the potential of a feedstock (processing costs and market values for the associated products).
- identify the required treatment for a feedstock.
- evaluate the quality of the products after the conversion process.

A second objective of the characterization is to determine the thermo physical property distributions representing the crude oil or petroleum fraction. These properties are used as inputs to thermodynamic phase behavior and physical property models. The two most commonly used characterization methods in refinery applications are based on distillation and solubility assays.

2.4.1 Distillation Curves and Boiling Point Distributions

Crude oils and petroleum streams are typically characterized starting from a fractionation of the petroleum into a number of components and pseudo-components that represent boiling point intervals. True boiling point (ASTM Standard D2892 2009) distillation is the most accurate method to determine boiling point intervals for a petroleum stream.

The True Boiling Point (TBP) refers to a distillation performed at high reflux ratio (≥5) and a large number of theoretical stages (≥14). Under these operation conditions, TBP distillation achieves a sharp separation of the components making up the crude oil. Therefore, the distribution of boiling points obtained with the TBP distillation describes the distribution of the components or pseudo-components in the crude oil mixture in order of boiling point.

However, TBP is not practical for most applications because the assay is time consuming and costly. TBP is instead approximated from other types of assays using empirical correlations. Examples of boiling point assays are ASTM standards D86, D1160 and D 5236. TBP can also be estimated from chromatographic analysis such as simulated distillation. Recently, a Spinning Band Distillation (SBD) method has been adapted to measure the boiling point distribution of crude oils equivalent to TBP. SBD has a large number of theoretical stages, depending on the length of the band in the column, and it can be performed at both atmospheric and vacuum pressures at any desired reflux ratio. It takes shorter operation time and requires less sample than the standard TBP ASTM method. All of these distillation methods are briefly presented in When the crude oil is characterized by any of the distillation methods except TBP, the distillation data must be converted to TBP. When the distillation is performed under vacuum, the temperature data also must be converted to atmospheric equivalent temperature. API has published standard procedures to interconvert data obtained from tests such as D86 and D1160 to an equivalent TBP curve, including the Riazi-Daubert method and the Maxwell and Bonnell method (ASTM Standard D5236, 2007; Riazi, 2005).

Note that the temperature measurements in most of the distillation methods, with the exception of ASTM 5236, do not represent a true thermodynamic state and the interconversion methods are empirically based correlations with no thermodynamic meaning. Therefore, inconsistencies are sometimes unavoidable.

Table2.3: Description of different methods to determine boiling point distributions

Method	Description
ASTM D2892(TBP)	Vacuum distillation. Sharp separation of component by their boiling point. Uses 14 to 18 theoretical stages and reflux ratio of 5:1
ASTM D5236	An extension of ASTM D2892. Allows measurements of boiling point of the mixture at reduced pressure.
ASTM D86	Atmospheric distillation. Limited by the cracking temperature of the crude oil samples (~344°c). D86 is not practical for heavy oil samples. Temperature are measured at the vapor phase.
ASTM D1160	Vacuum distillation for high boiling samples. Reduced pressure to avoid cracking. Samples can be distilled up to ~538°c. however, it is inadequate for heavy oil and bitumen.
Advanced distillation curve(Bruno 2006)	Atmospheric distillation. It is an improved setup of ASTM D86 that allows accurate measurements of real/instant boiling point of the mixture and provides accurate volume measurements.
Spinning Band Distillation(Adapted from ASTM D2892)	Atmospheric or vacuum distillation. The fractionation device is a spinning band which rotates at high speed giving efficiencies up to 50 theoretical plates and allowing high reflux ratio. Allows collection of fractions at desired pressure (>1 mmHg) and temperature intervals (Maximum pot temperature of 300°c to avoid cracking).
ASTM D3710,D2887,extended D2887(simDist,C30+,or SCN)	Gas chromatography assays used to predict boiling point distribution. The standard is usually a mixture of normal paraffins. Some “customized’’ aromatic corrections have been developed.

A new distillation method was developed by (Bruno, 2006) at (NIST 2014) to measure thermodynamic boiling points of conventional oils and also have more accurate volume measurements. This method is known as the Advanced Distillation Curve (ADC) or composition-explicit approach. ADC distillation can be also used to medium oils and is being adapted to work under conventional vacuum for heavier oils.

Boiling point distributions have limited applicability for heavy oils and bitumen because they contain high molecular mass fractions that decompose before they boil with conventional vacuum. Heavy oil and bitumen contain up to 70wt% of “non-distillable” components with conventional vacuum. Recently, (Castellanos-Diaz *et al.* 2014) developed a static Deep Vacuum Laboratory Apparatus (DVLA) to measure saturation pressures of petroleum fractions from extra-heavy oils and bitumen. DVLA is currently being adapted to fractionate bitumen by volatility to extend the data from conventional vacuum distillation (Sánchez-Lemus *et al*., 2014).

2.4.2 Crude Oil Pseudo-Components from Boiling Point Distributions

In order to use the TBP data for material balances and phase behavior modeling, the TBP curve must be split into a number of components and pseudo-components to represent the distribution of properties in the mixture. Generally, two techniques are applied to split the distributions:

- the pseudo-component approach.
- the continuous mixture approach (Riazi, 2005).

In the pseudo-component approach, the TBP curve is divided into

a number of sub fractions with known boiling point, specific gravity, and molecular weight (Tb, SG, and MW*)*. This procedure is also known as pseudoization. Note that, each fraction or pseudo-component is actually a mixture of many pure components that happen to boil at the same normal boiling point (or boiling point range). Therefore, each fraction may be further split into three (or more) pseudo-components to differentiate between chemical families such as paraffins, naphthenes, and aromatics if needed. Each fraction is assigned a unique set of properties such as density, molecular weight, critical properties, and elemental composition (Hay *et al.* 2013; Riazi, 2005).

In the continuous mixture approach, a Probability Distribution Function (PDF) is defined to describe the composition of the mixture. The PDF can be defined in terms of a measurable property, such as boiling point, molecular weight, refractive index, or carbon number, and varies from the value of the lightest to the heaviest component present in the mixture. When a mixture is represented with a PDF, it must eventually be presented in terms of pseudo-components (pseudoization of the distribution function). Therefore, both the pseudo-component and the continuous mixture approaches are used simultaneously. Both approaches work well when modeling physical operations in which the volatility is the major property of interest and no reactions and/or complex association are involved.

2.4.3 Property Estimation and Prediction for TBP Crude Oil Pseudo-Components:

Once the pseudo-components are defined, their properties must be measured or calculated. Typically, each pseudo-component from a boiling point distribution represents the average boiling temperature of the given interval on the boiling curve. The specific gravity and average molecular weight for each pseudo-component are determined from correlations. There are several correlations to estimate specific gravity based on Riazi-Daubert method using boiling point, molecular weight, or kinematic viscosity; however, the specific gravity of hydrocarbons and petroleum fractions is normally available from experimental data because it is easily measurable (Riazi, 2005).

Riazi-Daubert methods can be used to estimate the molecular weight of hydrocarbons with molecular weights up to 700 g/mol, using the constants recommended by (Tsonopoulos *et al.*, 1986) as follows

$$MW = 42.965\left[\exp\left(0.0002097 * T_b - 7.78712 * SG + 0.00208476 * T_b\right)\right]T_b^{1.26007} SG^{4.98308} \rightarrow 2.1$$

where **MW** is molecular weight, **T$_b$** is the boiling point and **SG** is the specific gravity.

This equation is recognized as the standard method of estimating molecular weight of petroleum fractions in the industry. Other characterization parameters can be measured for the bulk crude oil and used to constrain the correlations, particularly the average specific gravity and average molecular weight.

Once the **T$_b$**, **MW**, and **SG** are defined for each pseudo-component, other thermodynamic properties needed for phase behavior modeling, such as critical temperature, critical pressure and acentric factor, are then determined using correlations. There are a number of property correlations in the open literature and the most commonly used are the Lee-Kesler, Twu, and Riazi-Daubert methods.

For example, the critical temperature (**T$_c$**) and critical pressure (**P$_c$**) from the Riazi-Daubert method (Riazi, 2005) are given by:

$$T_c = 9.5233\left[\exp\left(-0.0009314 T_b - 0.544442 SG + 0.00064791 T_b SG\right)\right]T_b^{-0.81067} SG^{0.53691} \rightarrow 2.2$$

$$P_c = 3.1958 * 10^5\left[\exp\left(-0.008505 T_b - 4.8014 SG + 0.005749 T_b SG\right)\right]T_b^{-0.4844} SG^{4.0846} \rightarrow 2.3$$

These correlations were also adopted by the API and have been commonly used for industrial applications such as process simulators (Riazi,2005). Another useful property, particularly when modeling asphaltene precipitation, is the solubility parameter. (Barton,1991; Hansen, 2007) provided a table with solubility parameter data for several pure components. A correlation is required when the solubility parameter is unknown. The solubility parameter is related to the enthalpy of vaporization by definition as follows,

$$\delta_T = \left(\frac{\Delta H_{V_T} - RT}{V_T} \right)^{0.5} \xrightarrow{\hspace{8em}} 2.4$$

where **δ** is the solubility parameter, ΔHv_T is the enthalpy of vaporization at temperature T**,** R is the universal gas constant, and **v** is the molar volume. Measured data for enthalpy of vaporization are available in the literature for pure components. There are also several correlations in the open literature to estimate enthalpy of vaporization at a given temperature and pressure for pure components or pseudo-components with unknown properties. Some of the correlations are based on vapor pressure correlations derived from clausius-clapeyron equation and/or EOS and others are empirically developed. The enthalpy of vaporization at the boiling point, ΔH_{vb}, can be determined from the simplified Trouton's rule, Chen, or Vetere methods as shown below.

Chen Method (Poling *et al.***,** 2001):

$$\Delta H_{Vb} = RT_c T_{br} \frac{3.978 T_{br} - 3.958 + 1.555 \ln P_c}{1.07 - T_{br}} \xrightarrow{\hspace{8em}} 2.5$$

where T_{br} is the reduced temperature at the boiling point (Temperature in K and pressure in bars).

Vetere Method (Poling *et al.*, 2001):

When T_c and P_c are available (Temperature in K and pressure in bars):

$$\Delta H_{Vb} = RT_b \frac{(1 - T_{br})^{0.38} \left(\ln P_c - 0.513 + 0.5066 / \left(P_c T_{br}^2 \right) \right)}{1 - T_{br} + \left(1 - (1 - T_{br})^{0.38} \right) \ln T_{br}} (Vetere1) \rightarrow 2.6$$

and when T_c and P_c are not available:

$$\Delta H_{Vb} = RT_b \left(A + B \ln T_b + \frac{C T_b^{1.72}}{MW'} \right) (Vetere2) \cdot \xrightarrow{\hspace{8em}} 2.7$$

where A, B and C are constants. For hydrocarbons, A=3.298, B=1.015, C=0.00352, and MW′ is a “fictitious” molecular weight that is equal to the true molecular weight for most compounds.

Trouton’s Rule (Modified) (Hildebrand and Scott, 1964)

$$\Delta H_{Vb} = 87.5 * T_b \longrightarrow 2.8$$

Sometimes the enthalpy of vaporization is required at standard conditions and a temperature correction must be applied to ΔH_{vb}. A widely used correlation between ΔH_{vb} and temperature is the Watson correlation (Poling *et al.***,** 2001) given by:

$$\Delta H_{V2} = \Delta H_{V1}\left(\frac{1-T_{r2}}{1-T_{r1}}\right)^{n} \longrightarrow 2.9$$

where T_r is the reduced temperature, the subscripts 1 and 2 refer to Temperatures 1 and 2, and n is a constant. Common values for n are 0.375 or 0.38 (Poling *et al.***,** 2001; Weissberger, 1959). n can be also obtained with correlations (Viswanath and Kullor, 1965).

2.4.4 Solubility Based Methods for Crude Oil Characterization

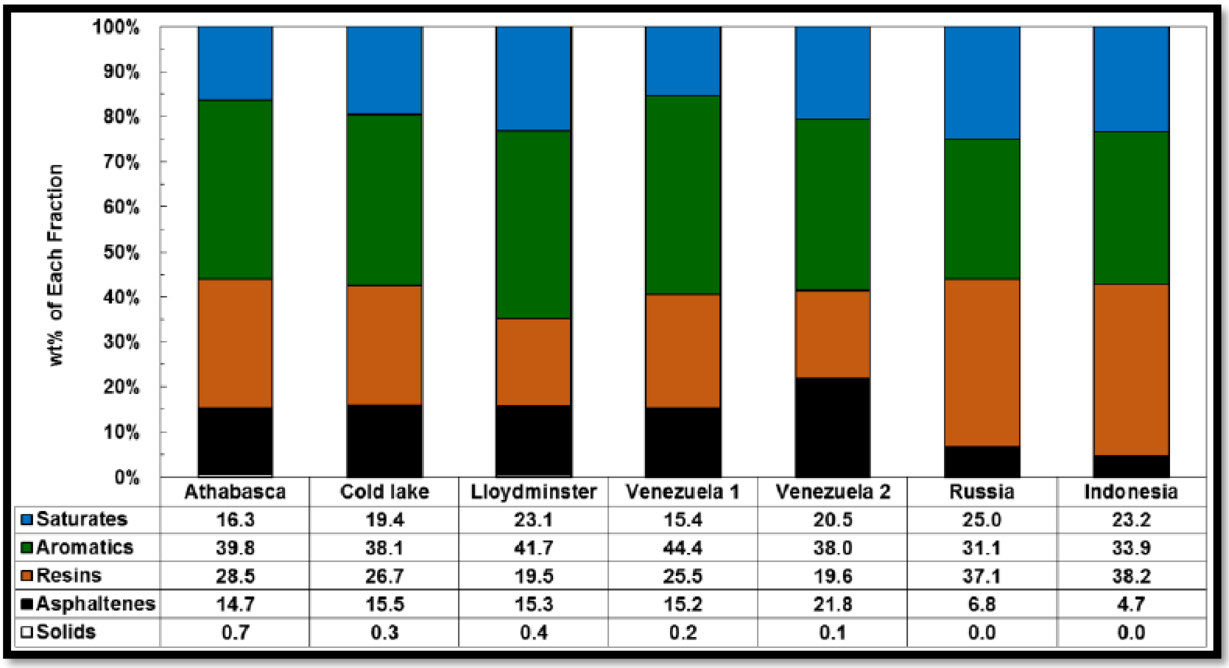
The boiling point characterization approach usually fails when used to model complex processes involving chemical reactions or extractive separations based on polarity (Hay *et al.*, 2013) as well as asphaltene precipitation from heavy oils. A n-Paraffin, Iso-paraffin, Olefin, Naphthene and Aromatic (PIONA) characterization technique has been developed by (Hay *et al.*, 2013) to capture the chemistry of a typical oil and was successfully used to track and estimate hydrocarbon mixture thermodynamic properties to be applied in chemical reaction processes in the refinery.

Other physical fractionation methods are also available to characterized heavy oils and bitumen based in solubility separation; for example, SARA analysis (adapted ASTM D4124 and ASTM D2007) where the oil is divided into four solubility and adsorption classes: Saturates Aromatics Resins and Asphaltenes (SARA). One advantage of the SARA fractionation approach is that separation between chemical families is achieved allowing for a better description of the solubility parameter distribution used for asphaltene precipitation modeling. There are also modified procedures to the solubility or adsorption type separations to separate polar or acidic sub fractions.

There are three general ASTM standard for separation of feedstocks into four or five pseudo-components, ASTM 2006, ASTM D2007, ASTM D4121. These methods use *n*-pentane as the solvent for the initial fractionation into asphaltenes and deasphalted oil; some have been modified to use *n*-heptane. Each assay separates the deasphaltened oil into different fractions: ASTM D2006 into polar compounds, 1st acidic paraffins, 2nd acidic paraffins, and saturates; ASTM D2007 into resins, aromatics, and saturates; ASTM D4124 into polar aromatics (resins), naphthenic-aromatics (aromatics), and saturates (Speight, 1998).

2.4.5 Crude Oil Pseudo-Components from SARA Fractionation and their Properties

A commonly used method to characterized heavy oils and bitumen is SARA fractionation in which the oil is separated into saturates, aromatics, resins and asphaltenes depending on their solubility and polarity. Figure 2.2 shows typical SARA composition of heavy oils from different sources (Diana,2014).



	Athabasca	Cold lake	Lloydminster	Venezuela 1	Venezuela 2	Russia	Indonesia
🟦 Saturates	16.3	19.4	23.1	15.4	20.5	25.0	23.2
🟩 Aromatics	39.8	38.1	41.7	44.4	38.0	31.1	33.9
🟧 Resins	28.5	26.7	19.5	25.5	19.6	37.1	38.2
■ Asphaltenes	14.7	15.5	15.3	15.2	21.8	6.8	4.7
□ Solids	0.7	0.3	0.4	0.2	0.1	0.0	0.0

Figure 2.2: SARA fraction of crude and heavy oils from different sources (Data from Akbarzadeh, et al .2005a)

2.5 Deasphalting and gasification process

2.5.1 Introduction

Upgrading heavy oil – be it heavy crude oil in the oil field or heavy bottoms streams in the refinery – is an increasingly prevalent means of extracting maximum value from each barrel of oil produced. Upgrading can convert marginal heavy crude oil into light, higher value crude, and can convert heavy, sour refinery bottoms into valuable transportation fuels. On the downside, most upgrading techniques leave behind an even heavier residue whose disposition costs may approach the value of the upgrade itself.

Solvent deasphalting is used in refineries processing heavy crude oils to upgrade heavy bottoms streams to deasphalted oil that may be processed to produce transportation fuels.

The process may also be used in the oil field to enhance the value of heavy crude oil before it gets to the refinery. The byproduct of deasphalting is an asphaltene stream for which it is often difficult to find a beneficial use.

By integrating heavy oil deasphalting design with that of gasification, important synergies may be realized. These include reductions in capital cost, increases in energy efficiency, and enhanced performance and profitability of each unit. Environmental benefits are also obtained. In fact, the combined deasphalting gasification unit may provide benefits that economically justify the combined processes in instances where neither of the technologies is justifiable on its own.

The Texaco Gasification Process is a market leader in the conversion of heavy oils, petroleum coke, and other heavy petroleum streams, to valuable products. The TGP is often employed as a means to beneficially use lower value streams that are produced by existing units. In this case, the design and operation of the bottoms processing unit and the TGP unit are done independently.

The TGP allows conversion of hydrocarbons, including asphaltenes, into synthesis gas that consists primarily of hydrogen, carbon monoxide, carbon dioxide, and water.

The Texaco Gasification Power System technology combines the TGP with combined-cycle power and cogeneration technology to produce steam and power. The Texaco Hydrogen Generation Process combines TGP technology with that of hydrogen production. The syngas may also be used as a precursor of other varied chemicals such as methanol, ammonia, and Oxo chemicals.

2.5.2 Technology Applications

An integrated solvent deasphalting/gasification facility is an attractive alternative for upgrading heavy oils economically. The technology can be utilized by refiners that process heavy crude oils and by oil field producers of heavy crude oils.

Refinery applications: For refiners, an integrated solvent deasphalting/gasification unit can increase the throughput or the crude flexibility of the refinery without creating a new, highly undesirable heavy oil stream. Typically, the addition of a solvent deasphalting unit to process vacuum tower bottoms increases a refinery's production of diesel oil. The deasphalted oil is converted to diesel using hydro treating and catalytic ("cat") cracking. Unfortunately, the deasphalter bottoms often need to be blended with product diesel oil to produce a viable outlet for these bottoms.

The Texaco Gasification Process (TGP) is capable of converting these deasphalter bottoms to synthesis gas ("syngas"). This syngas may be utilized by the refiner in a number of ways. The syngas can be converted to hydrogen by use of the Texaco Hydrogen Generation Process (THGP), which may be used in the refinery for hydrocracking and hydro treating. The syngas may also be used by Texaco Gasification Power Systems (TGPS) cogeneration facilities to provide low cost power and steam to the refinery. If the refinery is part of a petrochemical complex, the syngas can be used a chemical feed stock.

Oil fields: For heavy crude oil producers, the integrated solvent deasphalting/gasification unit can increase the value of their crude. Deasphalting removes the heavy components, reduces the metal content, reduces the Conradson carbon, and increases the API gravity of the crude. The lighter crude is more easily transported and has properties much closer to the design crude oils of most refineries. This allows the refinery to maximize refined product production allowing the refinery to justify a higher crude price.

The gasification unit provides the oil producer with syngas that can be handled in several ways. Most commonly, the syngas would feed a TGPS cogeneration unit to produce power and steam. The steam generated would then be used for well injection to enhance oil production in the field, and the power would be sold. The syngas also may be sold to third parties for its chemical value.

2.5.3 Process Descriptions

2.5.3.1 Deasphalting:

A diagram of a deasphalting unit is given as figure below.

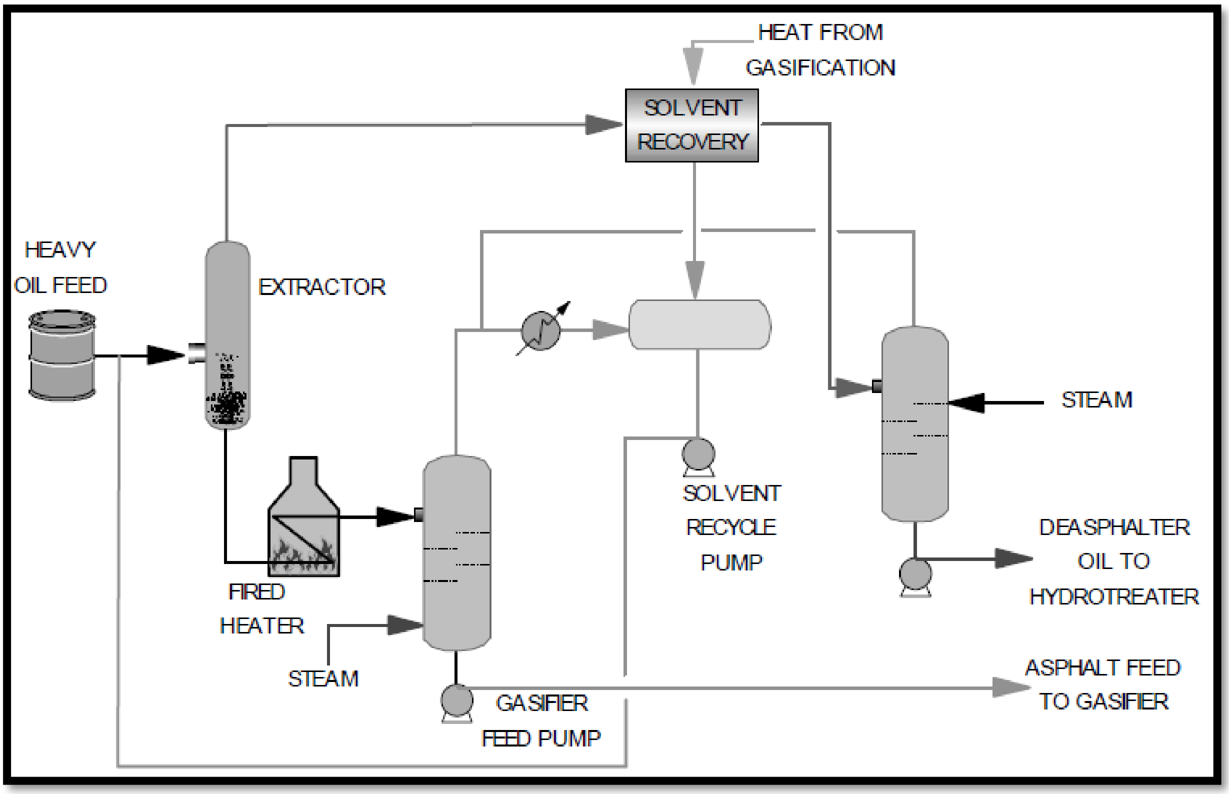


Figure2.3: Diagram of a deasphalting unit (Paul S. Wallace et al., 1998).

As crude oils are vacuum distilled, long chain paraffinic material and asphaltenes accumulate in the bottoms, or vacuum residue. As much as 80% of the residue from vacuum crude oil towers is paraffinic material that can be upgraded to diesel fuel.

The paraffinic components must be separated from the asphaltenes so that they can be cracked in conventional cracking units. This separation may be accomplished using solvent extraction. The extractor uses a hydrocarbon such as propane, butane or pentane to extract the paraffinic components from the feed stream. The heavy oil feed is mixed with the solvent. The asphaltenes are insoluble in the solvent and are separated from the paraffinic components by settling. The extractor produces solvent-rich De Asphalted Oil (DAO) and an asphaltene stream that contains some residual solvent.

The solvent-rich DAO is heated and flashed to recover the solvent. Alternatively, the solvent-rich DAO is heated to supercritical conditions and separated by settling. In either case, heat must be supplied to the process to achieve separation of the solvent from the DAO. Fired heaters and high-pressure steam are common sources for the heat. The solvent is returned to the extractor, and DAO is routed to a steam stripper for final solvent recovery. Typically, the DAO is then hydro-treated to remove sulfur and acids and to maximize yield in the downstream cracking units *(*Paul S. Wallace *et al*., 1998).

The solvent that is entrained in the asphaltenes must also be recovered. The solvent containing asphaltenes are heated above the minimum asphalt pumping temperature. This ensures that the asphalt will be pumpable after the solvent is removed. The heat source is typically a fired heater or high-pressure steam. The solvent is steam stripped from the asphaltenes in at rayed tower. The solvent is recycled to the extractor. The asphaltenes leave the stripper hot and must be cooled prior to blending for sales.

Solvent deasphalting is a cost-effective way to produce oil that can be converted to diesel from residual distillation products. The bottoms from the process are highly viscous at ambient temperature. To market this material, it must be “cut” blended with a significant amount of expensive distillate products. This requirement it is often detrimental to the unit’s overall economics.

The deasphalting unit requires significant amounts of heat to recover the solvent used in the extraction. Whether a fired heater generates this heat or it is obtained by the use of high-pressure steam, the energy cost is significant. Finally, when a fired heater is used, stack emissions result.

2.5.3.2 Texaco Gasification Process

The Texaco Gasification Process was developed in the late 1940s. It was intended to produce hydrogen and carbon monoxide - syngas - for chemical plant and refinery applications. It was designed to be fed natural gas. In the 1950s, it was modified for heavy oil feeds, in the 1970s for solid feeds like coal, and in the 1980s for petroleum coke.

Nearly from its inception, the process has been an attractive means for hydrogen production. The technology for this production has become the Texaco Hydrogen Generation Process (THGP). In the late 1970s, the process was modified to mate to a combined cycle power plant. This technology became specialized to the degree that it has become its own technology, now named Texaco Gasification Power Systems (TGPS).

Texaco gasifiers will soon produce 130248 standard cubic Meter of syngas per day in forty-nine currently operating installations around the world and nineteen more in engineering and construction phases. The majority of this capacity is still used for chemicals production, but the percentage used for power production has been rising the fastest. Soon at least 45% of the syngas generated by Texaco gasifiers will be used for power production.

Among commercially proven technologies, Texaco Gasification Process based plants remain the most environmentally benign means of generating valuable products from sulfur-containing feedstocks. Power plants with TGPS technology emit a fraction of the NO_x and SO_x pollutants that are produced from conventional or fluidized bed boiler installations. And even advanced boiler systems produce solid wastes in quantities far in excess of those produced in TGPS plants (Paul S. Wallace *et al.*, 1998).

Texaco gasification converts heavy oils such as vacuum residue and asphaltenes into synthesis gas (syngas) which is primarily hydrogen and carbon monoxide. Syngas has a variety of uses. Power, steam, hydrogen, and other products can be produced in any combination. To obtain maximum economic benefit from the unit, a low value feedstock is desirable. The heat generated by the gasification reaction is recovered as the product gas is cooled.

When the quench version of Texaco gasification is employed, the steam generated is of medium and low pressure. A quench gasification flow scheme as would be applied to the integration with deasphalting is shown in Figure 2.5.

Note that the low-level heat used for deasphalting integration is the last stage of syngas cooling. In non-integrated cases, much of this heat is uneconomical to recover and is lost to air fans and to cooling water exchangers.

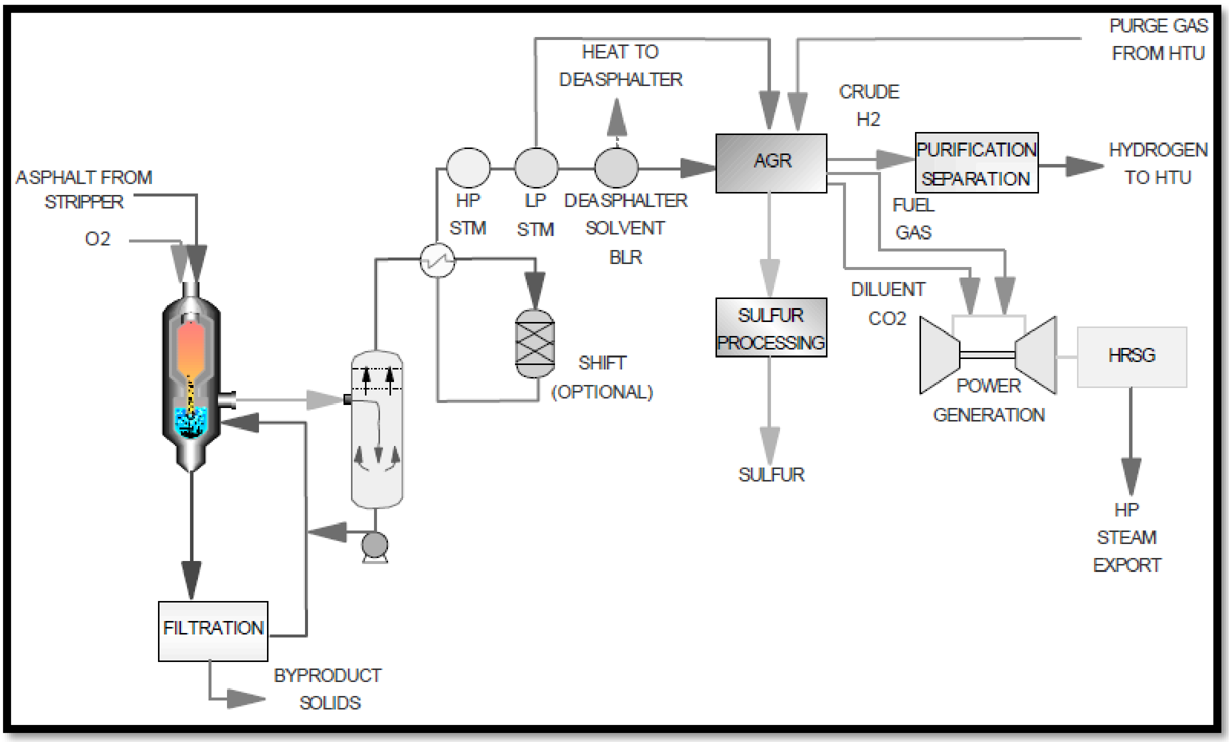


Figure2.4: Diagram of a gasification unit (Paul S. Wallace et al., 1998)

2.5.4 Integrated Gasification – Deasphalting:

A summary of the feeds and products of integrated deasphalting and gasification is given in Figure 2.5 Integrating solvent deasphalting with gasification enhances the operation and economics of both technologies.

There are three main synergies between the deasphalter and the gasifier:

- Beneficial use of the asphaltenes
- Internal consumption of low-level heat
- Production of hydrogen for DAO treating
- Recovery of hydrogen from hydro treating purge gas

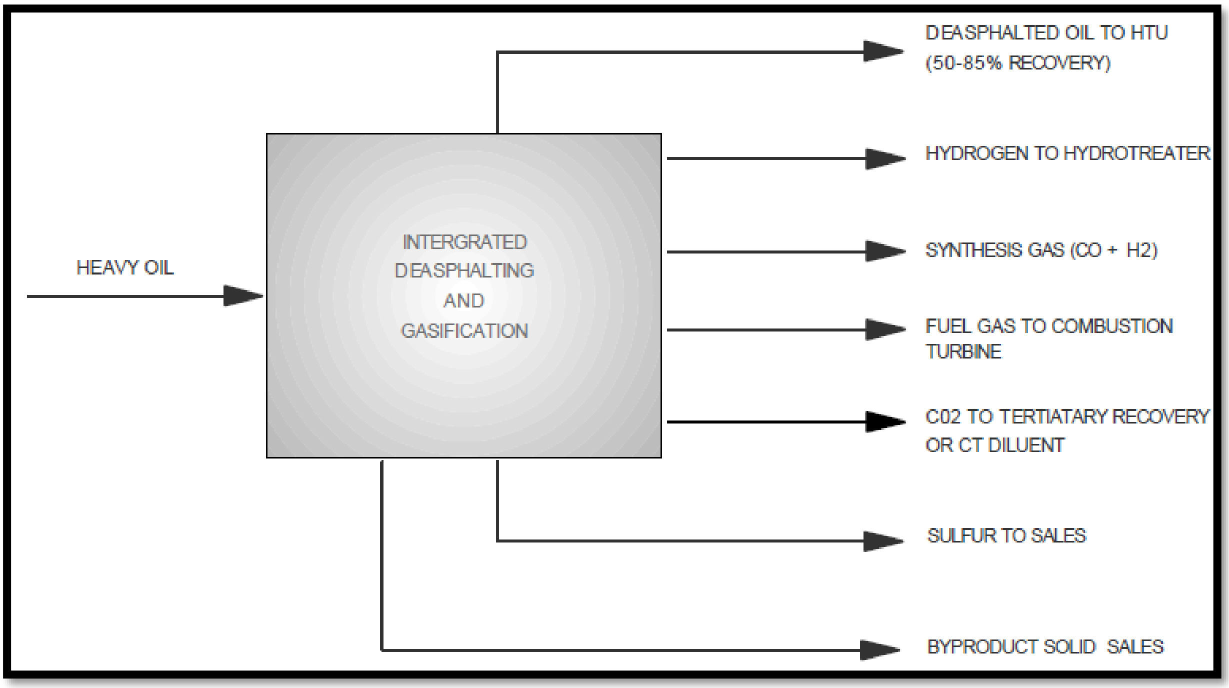


Figure2.5: Integrated deasphalting and gasification unit (Paul S. Wallace et al., 1998)

The bottoms from the deasphalter, the asphaltenes, may be gasified. This eliminates the need to use expensive distillate blending streams to make it marketable. The asphaltenes are a low value feedstock for gasification, which enhances its profitability. In nonintegrated deasphalters, however, the bottoms' viscosity may be high enough to make pumping them to the gasifier difficult. Heating the asphaltenes may be done to improve its pumping characteristics. Unfortunately, this material has poor heat transfer characteristics and heating it without coking is prohibitively expensive. Instead, the asphaltenes are typically cut with light oil to make them pumpable. This too is expensive.

The viscosity of the asphaltenes is higher when longer chain solvents such as pentane are used in the deasphalting process. Pentane solvent is used instead of shorter chain solvents such as propane or butane because they result in a "deeper cut", that is, a greater yield of deasphalted oil. The pentane asphaltenes are more concentrated, and require higher temperatures to maintain pump able viscosity.

With the integrated deasphalter-gasification unit, the gasifier feed is taken directly from the asphalt stripper. The asphalt is heated to the temperature required for optimal pumping to the gasifier prior to solvent removal, when its heat transfer characteristics are more favorable. The result is that viscosity limits on the asphaltenes are eliminated.

Most importantly, high severity deasphalting with pentane produces a higher yield of DAO and ultimately enhances the refinery's production of diesel oil.

The gasifier charge pump draws from the bottom of the stripper and routes the material directly to the gasifier. The working volume in the bottom of the stripper acts as a charge drum for the gasifier and minimizes the storage time for the asphaltenes. This short storage time eliminates the potential of the hot asphaltenes to polymerize.

If the deasphalter shuts down, the gasifier continues to operate using the heavy oil feed to the deasphalter. The feed to the deasphalter can be gasified with only minor adjustments to the operating parameters. The deasphalter feed is not as economically advantageous a feed as the deasphalter bottoms. However, it will allow the gasifier to remain operating during deasphalter outages.

A key synergy of integrated solvent deasphalting and gasification is the sharing of each process' heat. The solvent deasphalting process requires a significant amount of heating to recycle the solvent used in the asphaltene extraction. The heat is used to strip the solvent from the oil and the asphaltene streams so that it can be recovered and returned to the process. The gasification process produces heat that can be used for this solvent recovery in the deasphalting unit.

Integrating the solvent deasphalter and the gasification unit enhances the energy balance between both units. The low-level heat from quench gasification is used directly in a multi-stage sub-critical vaporization. No external source of heat is required to separate the solvent from the DAO. The mid-level heat from gasification is used to minimize stripping and asphalt fired heater duties.

The products of deasphalting and gasification can also be beneficially integrated. The De Asphalted Oil (DAO) requires hydro treating and cat cracking to become diesel.

Hydrogen is required for this treating, which is a primary product of gasification. The hydrogen can be generated from the asphaltene to eliminate the need for any externally supplied hydrogen.

The purge gas from hydro treating may be routed to the gasifier. Hydrogen is then recovered in the same process used to generate syngas-borne hydrogen. This eliminates losses of hydrogen to the fuel gas system. In addition, this allows the sulfur in the DAO to be captured internally without increasing the load on the refinery sulfur facilities.

Some of the advantages of integrated deasphalting and gasification may be realized by adding TGP units to existing solvent deasphalters. The gasification of the bottoms eliminates the need to blend the bottoms stream for sale. They can also generate hydrogen for the hydro-treating needs of the refiner. These benefits can be obtained without retrofitting the existing deasphalting unit. The benefits of heat integration may be added if the facility is willing to retrofit the deasphalter.

2.5.5 Integration Advantages

As indicated in the preceding paragraphs, there are numerous benefits obtained in integrating deasphalting and gasification:

The integration allows heavy oils to be upgraded economically.

The asphaltenes are gasified, converting this the lower value material to high value syngas.

Syngas can be used as a source for hydrogen, carbon monoxide, as a precursor of Oxo chemicals, or as a fuel in combustion turbines. This eliminates the need to use expensive distillate to blend off the heavy material. In addition, the gasification unit is able to supply most of the heat required by the deasphalting unit. The integration reduces capital and operating costs of both processes.

In conventional deasphalters, a fired heater or high-pressure steam from a boiler is typically used to produce the heat necessary for the deasphalting process. This represents a significant use of fuel. When the process heat available from the gasifier is used to heat the deasphalter streams, the capital and operating cost of deasphalting is reduced.

The capital cost of the integrated gasification/solvent deasphalter unit is lower due to shared equipment. In conventional deasphalters, heat exchangers or fired heater are required to provide energy for solvent separation. In the gasification unit, heat exchangers and air fan coolers are required to cool the syngas. These services are combined in the integrated unit, reducing the total number of exchangers required and the overall capital cost.

In addition, other heat exchange surface area typically found in conventional deasphalters may be eliminated. When high level energy is used to supply heat for solvent recovery, an effort is made to recover as much heat as possible so that operating costs can be minimized. When the heat for solvent separation is being supplied from waste heat from the gasification process, this heat recovery is unwarranted.

In the integrated solvent deasphalter and gasification process, the deasphalter maximizes the use of low level heat (below 176.7 °C) instead of using large amounts of high level heat: steam at (69.96 - 104.4) bar or fired heaters. The total heat consumption in the deasphalter is (20-40) % higher with the integrated scheme. However, the operating cost is lower, because most of the heat used by the deasphalter is supplied from the gasifier. The energy from gasification would otherwise be expelled to the atmosphere. The operating costs are reduced, because a minimal amount of fuel is required supply heat to the solvent recovery process.

The integration allows higher yields of DAO. Often the recovery of DAO is limited by a viscosity specification for the asphaltenes.

Without such a limit, the asphaltenes may require large amounts of high value light oil to produce a pump able stream or a marketable product. This requirement increases with increasing asphaltene viscosity, and may eventually seriously harm the unit's economics.

With the integrated flow scheme the asphalt viscosity need not be limited. This is due to the close coupling of the gasifier feed pump to the asphalt stripper and the additional upstream heating of the asphalt solvent mixture. Without this viscosity limit, the refinery may operate the deasphalter more severely, increasing their DAO yield and ultimately, increasing their diesel oil production. The integration results in lower emissions. Since the integration of the deasphalter and gasification units minimizes the amount of fuel that must be consumed to generate the heat required for solvent recovery, less NOx, and carbon dioxide are emitted.

Also, the asphaltene material produced in the extractor typically contains a significant amount of sulfur components. These sulfur components become part of the residual fuel oil when the asphaltenes are blended with distillates. The sulfur is emitted when the residual fuel oil is combusted. However, when the asphaltenes are gasified the sulfur is converted to hydrogen sulfide.

This is removed from the syngas using conventional acid gas absorption technology and converted to elemental sulfur. As a result, the ultimate emissions of sulfur oxides to the atmosphere are substantially reduced. An example of a typical reduction in sulfur content leaving the refinery is shown in Figure below.

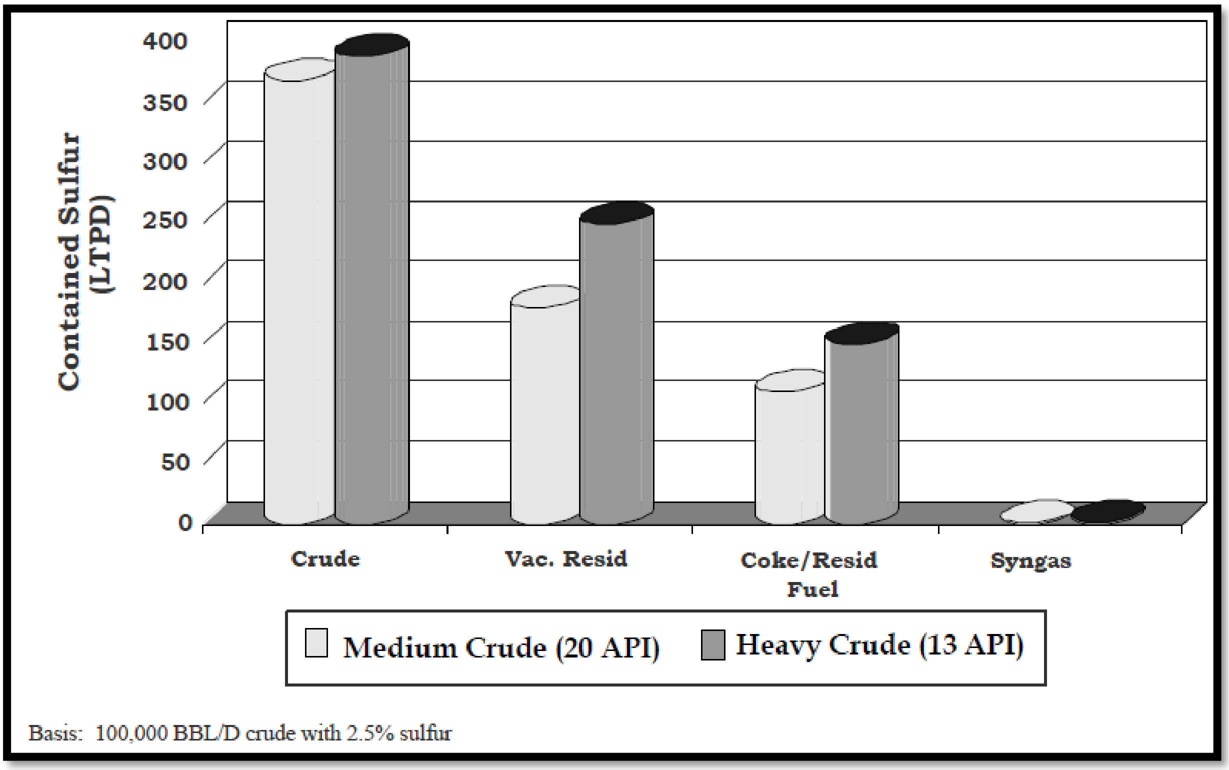


Figure 2.6: Sulfur content in the medium and heavy crude

*(*Paul S. Wallace *et al*., 1998)

At the same time, environmental concerns have increased, resulting in more rigorous specifications for petroleum products, including fuel oils. These trends have emphasized the importance of processes that convert the heavier oil fractions into lighter and more valuable clean products (Gray,1994).

CHAPTER THREE

Methodology

3.1 Introduction

Generally, in the field of scientific research it is important to describe in detail the methods which are followed up and used. In this work and in order to achieve the objectives of this thesis, the methods can be divided into three main categories as follows:

1. Data collection
2. Simulation program and Case study (EL Obeid refinery).
3. Addition a second atmospheric distillation unit and a Vacuum Distillation Unit(VDU) by Simulation Program Aspen Hysys.

3.2 Data Collection

Data of (Nile Blend, Rawat and Thargath) crudes was collected from EL Obeid Refinery Company and Central Petroleum Laboratory. The data consist of:

1. Historical data
2. Production data
3. Design data

3.2.1 Historical data

Crudes assay reports for (Nile blend, Rawat and Thargath). were collect The crude oil was characterized using the True Boiling Point (TBP) analysis (Table 3.1).

The assay report contains crude oil general tests (density, API, viscosity, pour point, water content, salt content, acid number, sulfur content, and asphaltenes results for crude); TBP distillation data; and the volume as well as weight percentage yield of crude distillation column products (Naphtha, Kerosene, Gas oil, and Residue). It also contains general test results for products.

3.2.2 Production Data

The weight percentage yield and the amount of crude oil products which were produced from the atmospheric distillation column were collected from El Obeid Refinery Company. The products consisted of Light and Heavy Naphtha, Kerosene, Gas oil and Atmospheric Residue.

Table 3.1: Nile Blend True Boiling Point Distillation Data

Cut Range °C	Yield %Weight	Cum. % Weight	Yield %Volume	Cum. % Volume
C4	0.06	0.06	0.09	0.09
C4 – 60	0.45	0.52	0.60	0.70
60 – 75	0.46	0.98	0.59	1.29
75 – 90	0.82	1.80	1.02	2.31
90 – 105	0.70	2.50	0.85	3.16
105 – 120	0.82	3.32	0.99	4.14
120 – 135	0.80	4.12	0.94	5.09
135 – 149	0.95	5.06	1.11	6.20
149 – 165	1.05	6.11	1.22	7.42
165 – 180	0.91	7.03	1.04	8.45
180 – 195	1.24	8.27	1.39	9.85
195 – 210	0.71	8.98	0.79	10.63
210 – 225	2.15	11.13	2.35	12.98
225 – 232	0.86	11.99	0.93	13.91
232 – 240	0.71	12.70	0.77	14.68
240 – 255	0.71	13.41	0.76	15.44
255 – 270	2.19	15.60	2.33	17.77
270 – 285	1.61	17.21	1.71	19.47
285 – 300	3.30	20.51	3.44	22.91
300 – 315	2.23	22.74	2.33	25.24
315 – 330	2.27	25.01	2.35	27.59
330 – 345	2.06	27.07	2.13	29.72
345 – 360	2.71	29.78	2.78	32.51
360 – 369	1.37	31.15	1.40	33.91
369 – 400	0.80	31.96	0.82	34.73
400 – 420	3.69	35.65	3.75	38.48
420 – 440	4.95	40.61	5.00	43.48
440 – 460	5.36	45.97	5.36	48.84
460 – 480	2.63	48.59	2.60	51.44
480 – 500	3.69	52.29	3.64	55.08

3.2.3 Design Data

A refinery process flow diagram was built using the following design data from HYSYS getting started module in order to build up the simulation.

Table 3.2: Heavy Naphtha Stripper

Tray no	1-4	
Tray diameter	87.3 cm	
Number of down comers per tray	1.0	
Down comer area	L	83.3 cm
	H	26.3 cm
Down comer weir height	4.5 cm	
Down comer clearance	3.2 cm	
Number of valves per tray	13	
Number of open valves	13	
Number of closed valves	0.0	
Valve open area	16.25 cm^2	
Opening area per tray	13x16.25 = 211.25 cm^2	
Distance between trays	45.5 cm	

Table 3.3: Gas Oil Stripper

Tray no	1-4	
Tray diameter	87.3 cm	
Number of down comers per tray	1.0	
Down comer area	L	83.3 cm
	H	26.3 cm
Down comer weir height	5.0 cm	
Down comer clearance	4.5 cm	
Number of valves per tray	13	
Number of open valves	13	
Number of closed valves	0.0	
Valve open area	16.25 cm^2	
Opening area per tray	13x16.25 = 211.25 cm^2	
Distance between trays	46.0 cm	

In any simulation problem, all feed streams must be completely defined before the calculation can begin. For a stream to be completely defined there must be enough information present to calculate the enthalpy. For mixtures, this requires the composition plus any two of the temperature, pressure, or vapor fraction.

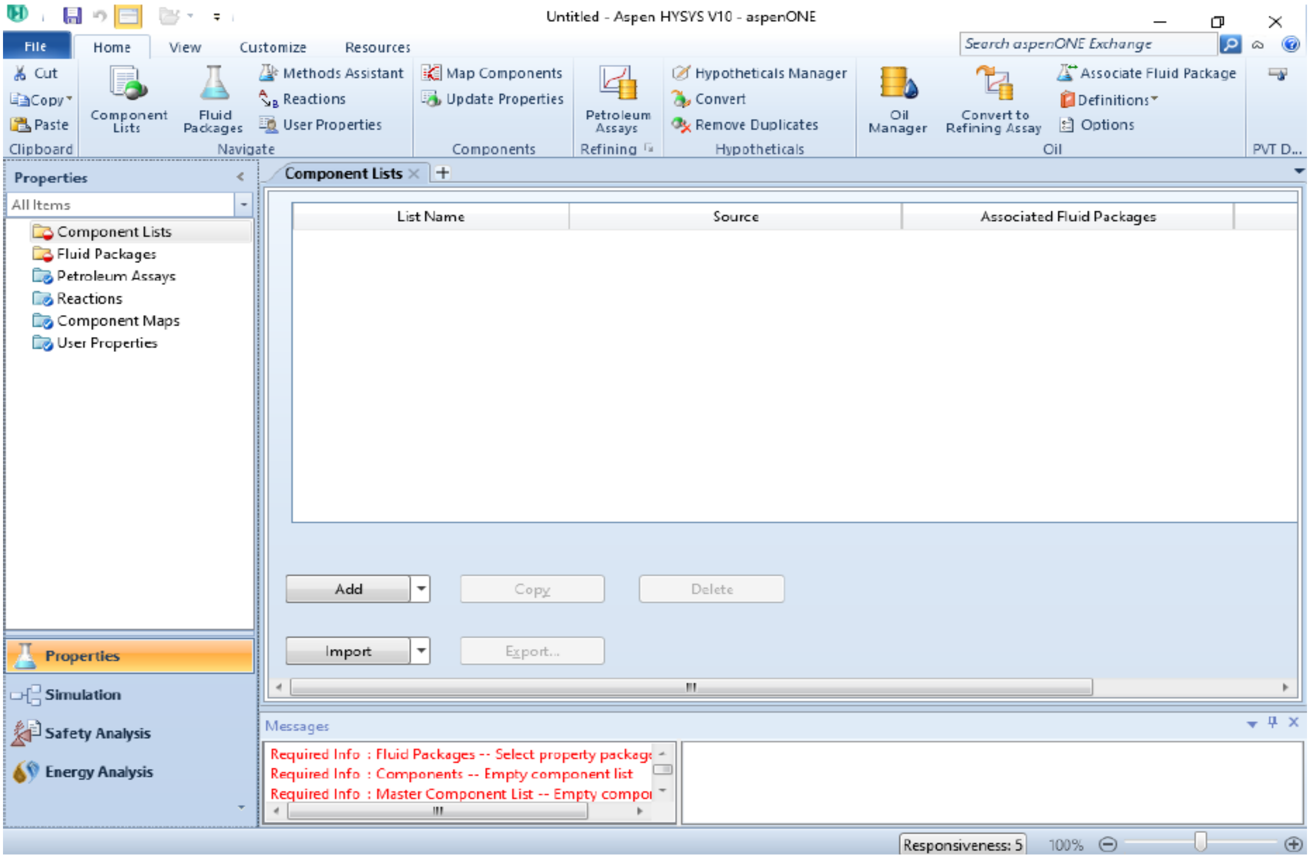
For pure components that are saturated or two phases, it is necessary to define the composition and the vapor fraction plus either the temperature or the pressure.

3.3 Simulation program

The Aspen Hysys (V10.0) simulation program was used for the simulation studies and the investigation of the effect of various operating parameters, because it offers a high degree of flexibility to accomplish a specific task. This flexibility combined with a consistent and logical approach to show these capabilities are delivered makes HYSYS an extremely adaptable process simulation tool (Technology, 2005).

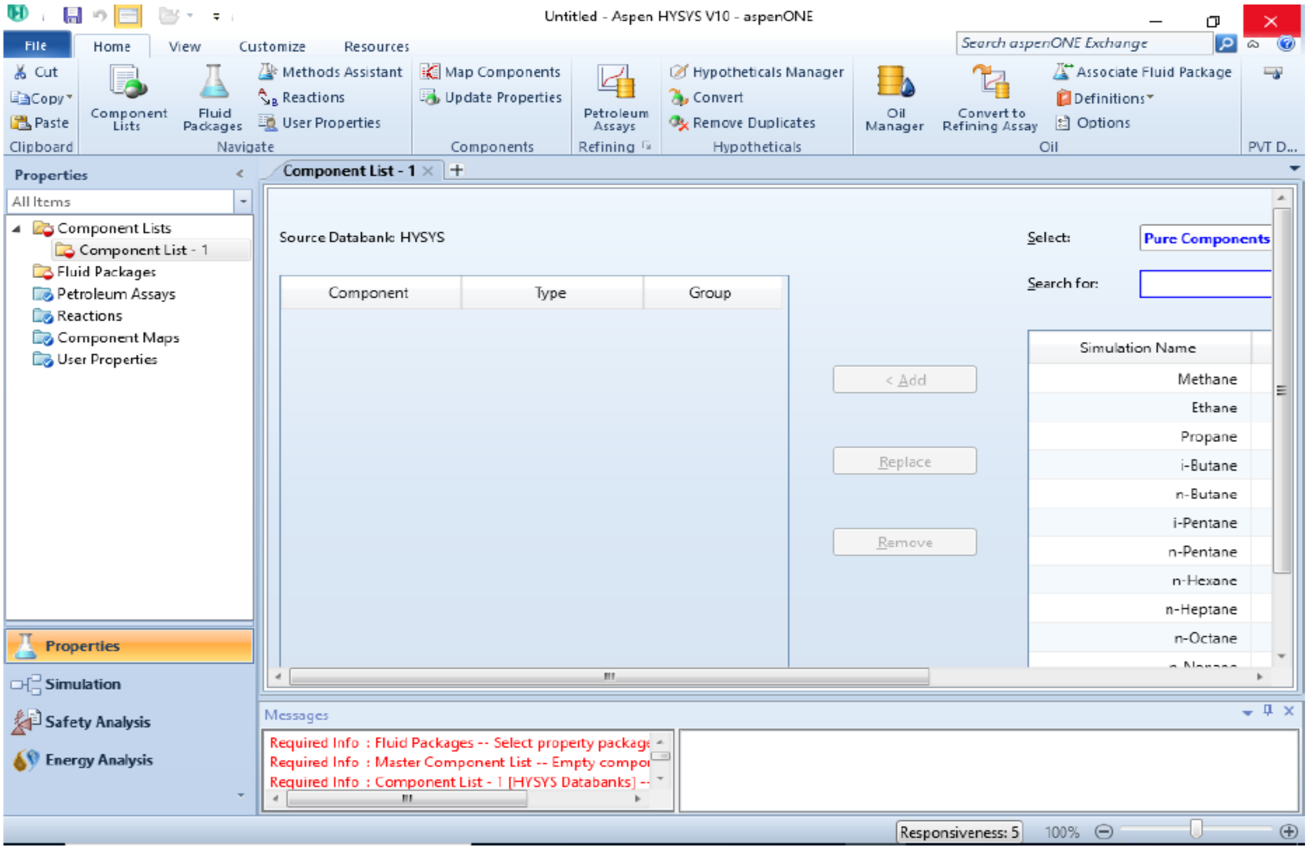
Start HYSYS and create a new case (Figure 3.1). The first thing which was done in order to build up the simulation case was selecting the components.

The component list for the fluid package were created, click the components lists (Figure 3.2), then click the add button. The component list property view appears. Add the following non-oil components: C_3, i-C_4, n-C_4, i-C_5, n-C_5 and H_2O (Figure 3.3).



Required Info : Fluid Packages -- Select property packag
Required Info : Components -- Empty component list
Required Info : Master Component List -- Empty compo

Figure 3.1: New case in Hysys program



Source Databank: HYSYS

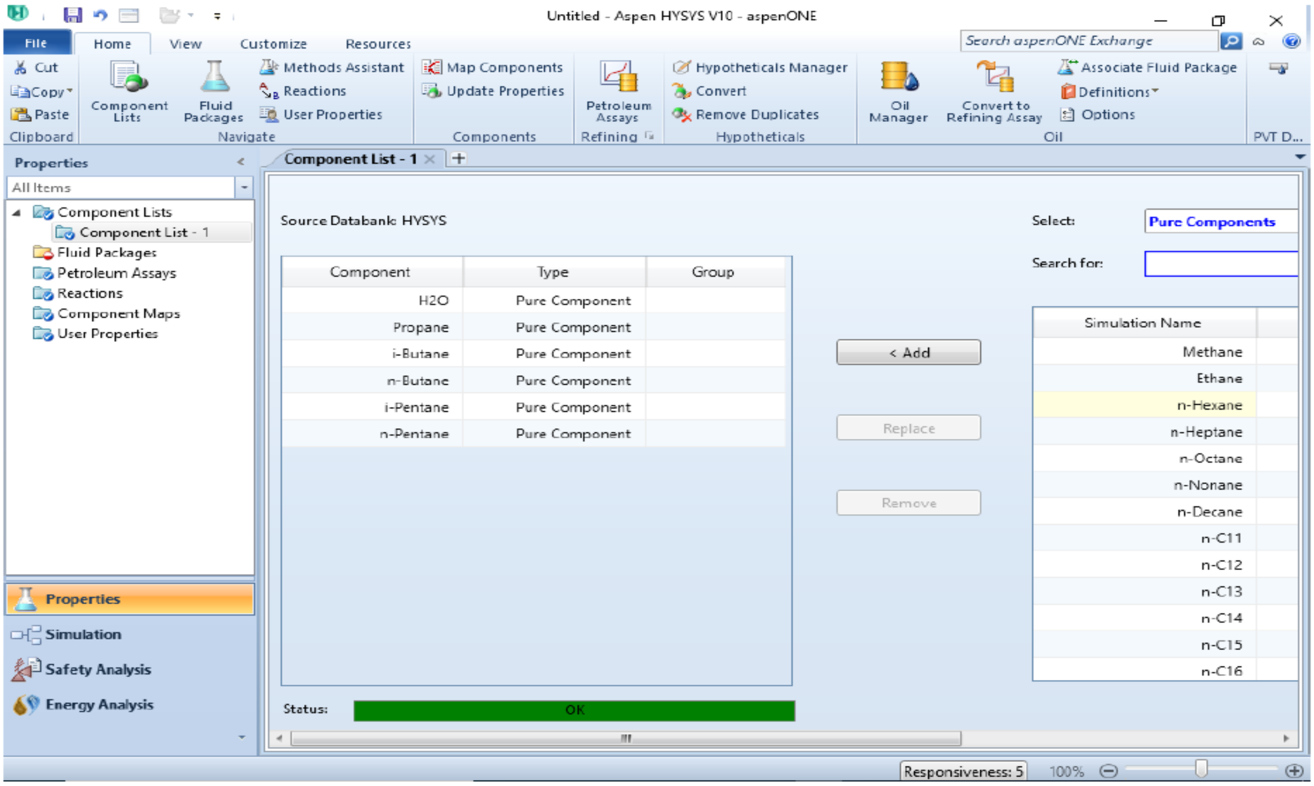
Select: Pure Components

Search for:

Simulation Name
Methane
Ethane
Propane
i-Butane
n-Butane
i-Pentane
n-Pentane
n-Hexane
n-Heptane
n-Octane

Required Info : Fluid Packages -- Select property packag
Required Info : Master Component List -- Empty compo
Required Info : Component List - 1 [HYSYS Databanks] --

Figure 3.2: Component list



Source Databank: HYSYS

Select: Pure Components

Search for:

Component	Type	Group
H2O	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	

Simulation Name
Methane
Ethane
n-Hexane
n-Heptane
n-Octane
n-Nonane
n-Decane
n-C11
n-C12
n-C13
n-C14
n-C15
n-C16

Status: OK

Figure 3.3: Components Selection

HYSYS has strong thermodynamic foundation. The inherent flexibility contributed through its design, combined with the unparalleled accuracy and robustness provided by its property package calculations lead to the representation of a more realistic model. In the simulation basis manager property view, the next task is to define a fluid package (Figure 3.4).

A fluid package contains the components and property methods used in calculations for a particular flowsheet. The equation of state models was chosen as a fluid property package, because it has proven to be very reliable in predicting the properties of most hydrocarbon based fluids over a wide range of operating conditions.

The Peng-Robinson equation of state was used among the equation of states models (Figure 3.6) because it has a high accuracy for a variety of systems over a wide range of conditions. It also rigorously solves most single-phase, two-phase, and three-phase systems with a high degree of efficiency and reliability (Technology, 2005a).

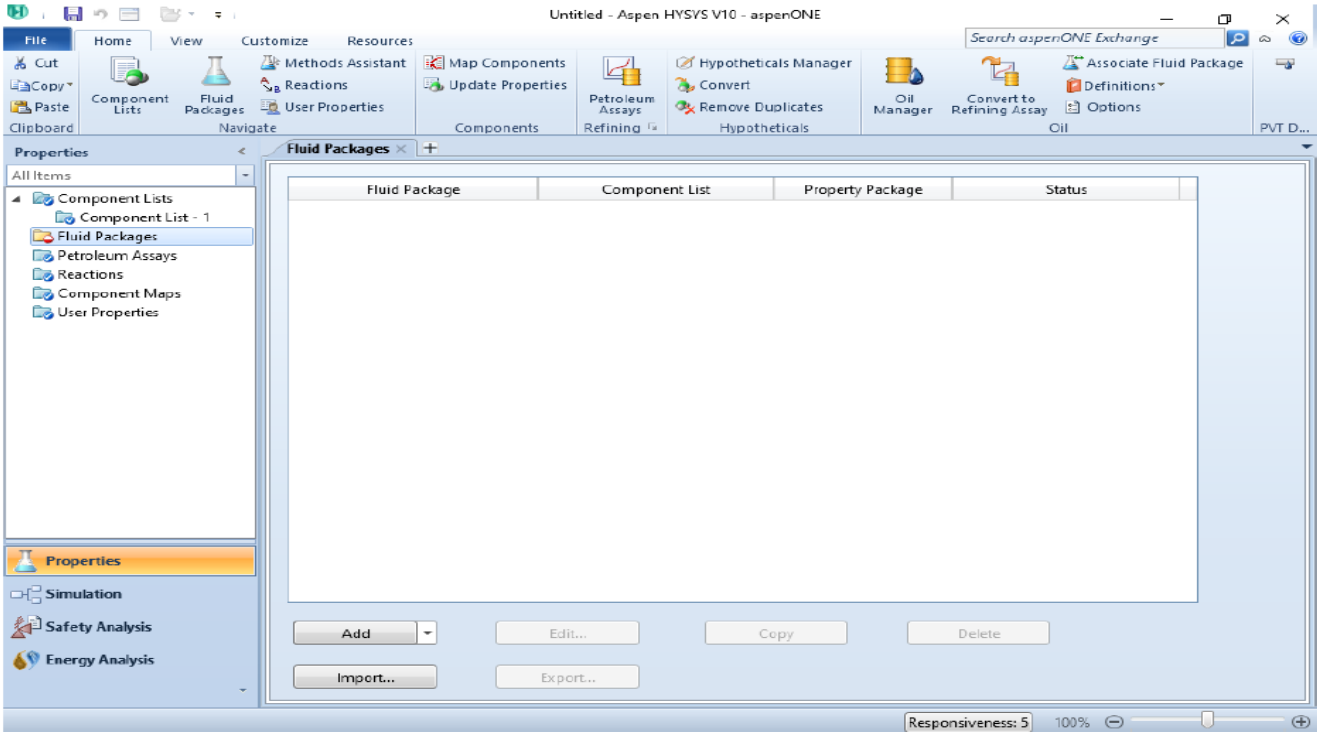
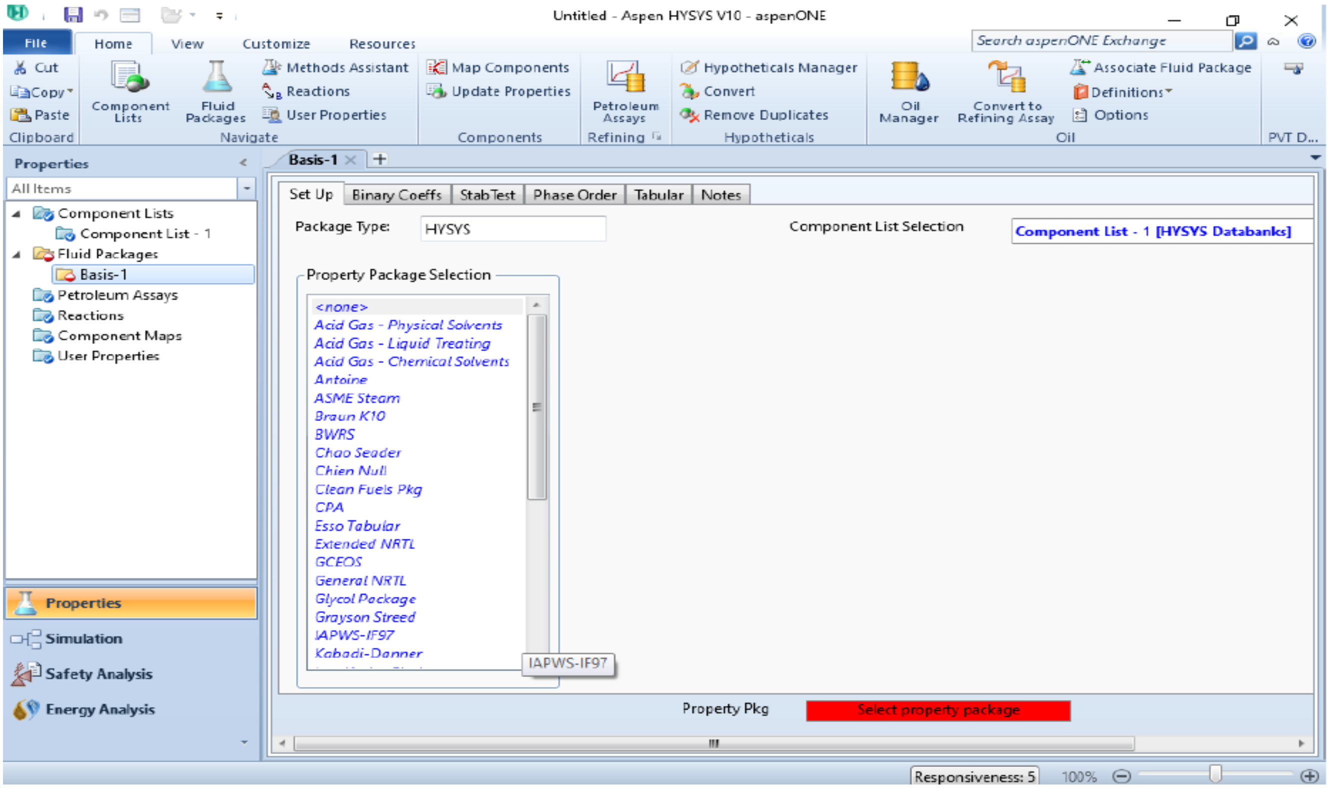


Figure 3.4: Fluid package selection

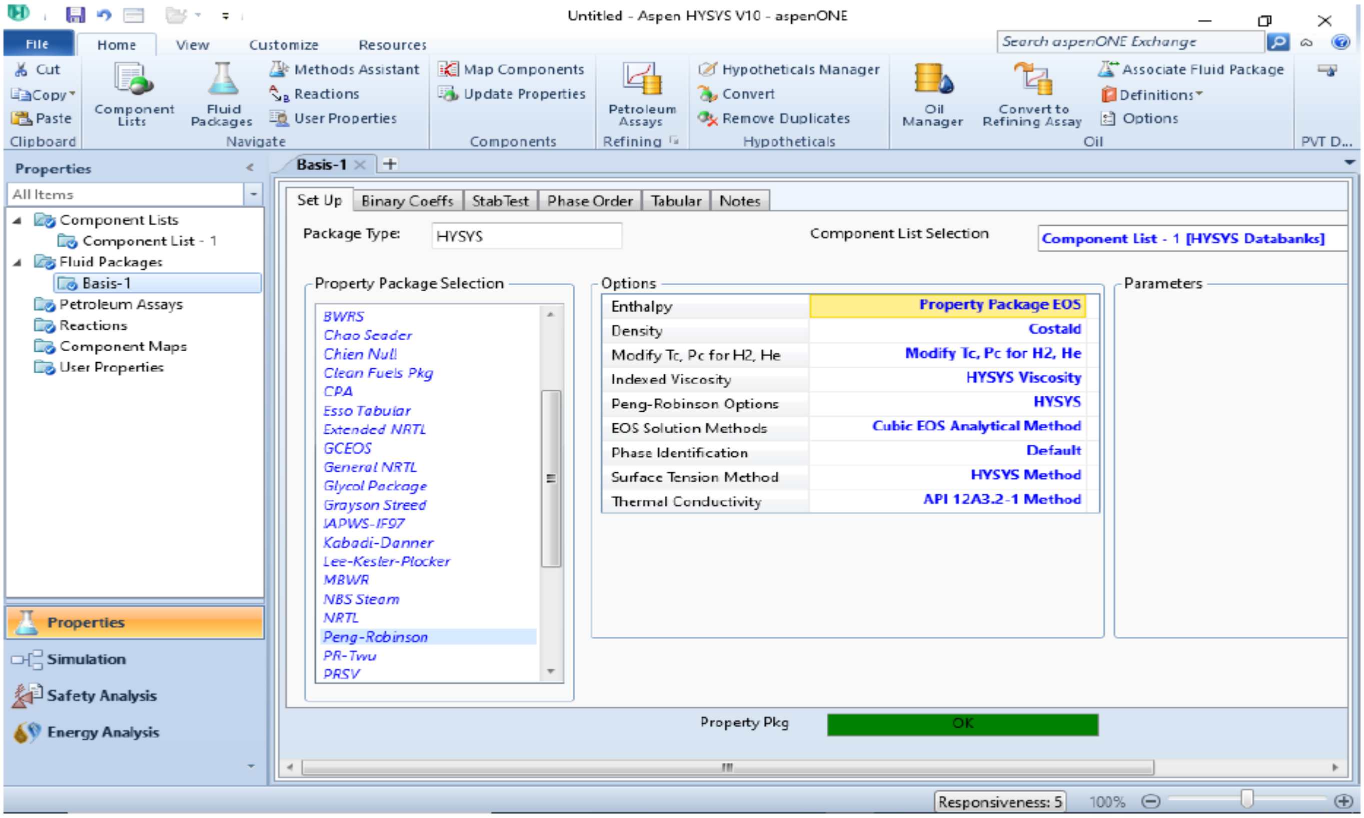


Package Type: HYSYS

Component List Selection: Component List - 1 [HYSYS Databanks]

Property Pkg: Select property package

Figure 3.5: Property package selection



Package Type: HYSYS

Component List Selection: Component List - 1 [HYSYS Databanks]

Options	
Enthalpy	Property Package EOS
Density	Costald
Modify Tc, Pc for H2, He	Modify Tc, Pc for H2, He
Indexed Viscosity	HYSYS Viscosity
Peng-Robinson Options	HYSYS
EOS Solution Methods	Cubic EOS Analytical Method
Phase Identification	Default
Surface Tension Method	HYSYS Method
Thermal Conductivity	API 12A3.2-1 Method

Property Pkg: OK

Figure 3.6: Selection of Peng-Robinson from equations of state

The Hysys oil characterization procedure is used to convert the laboratory data into petroleum hypocomponents. The oil characterization property view allows to create, modify, and otherwise manipulate the assays and blends for simulation case. In general, three steps must be completed while characterizing a crude oil:

1. Supply data to define the assay.
2. Cut the assay into hypothetical components by creating a blend.
3. Install the hypothetical components into the fluid package

after that oil manager was selected from toolbar (Figure 3.7), on the Assay tab, click the add button to create and view a new assay. The assay property view appears. When the property view for a new assay is opened, the property view contains minimal information. Depending on the assay data type, the property view is modified appropriately.

For this experiment, the assay is defined based on TBP data.

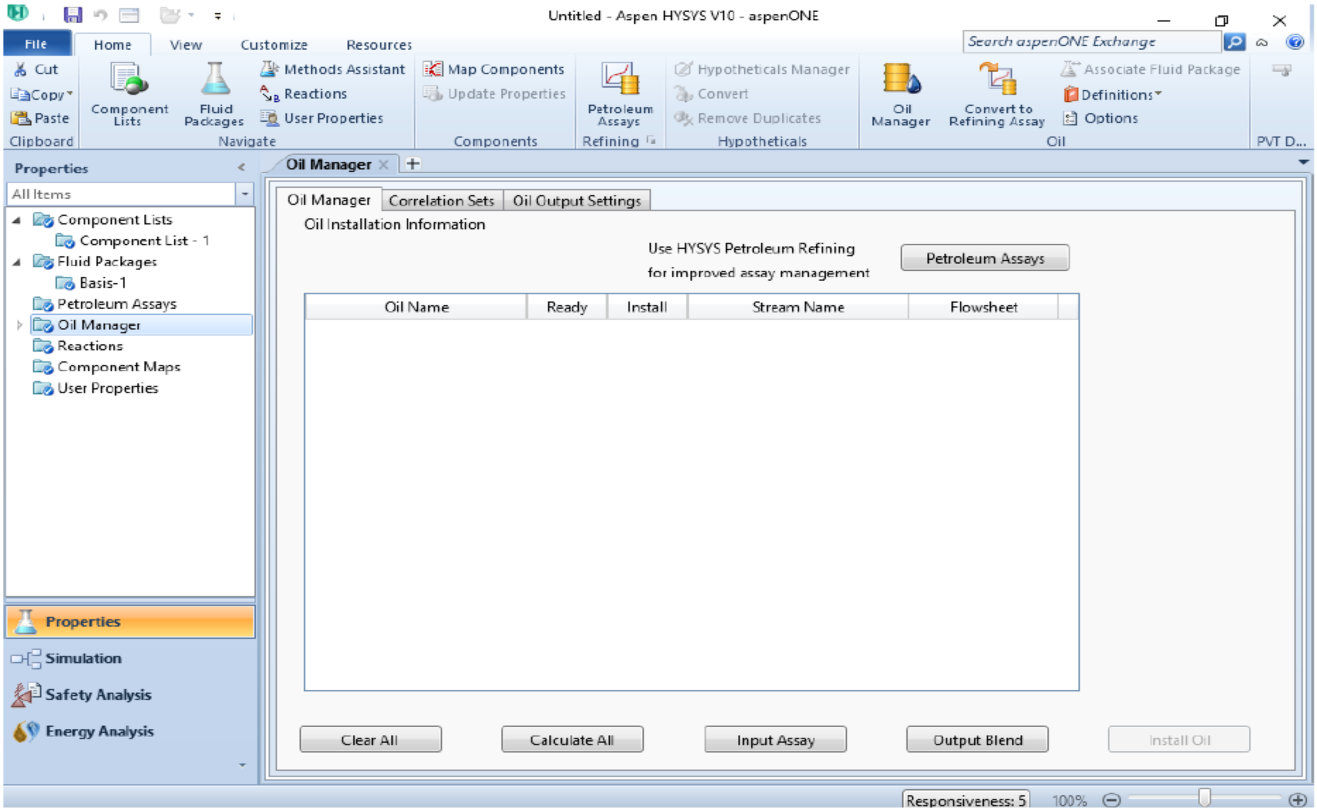
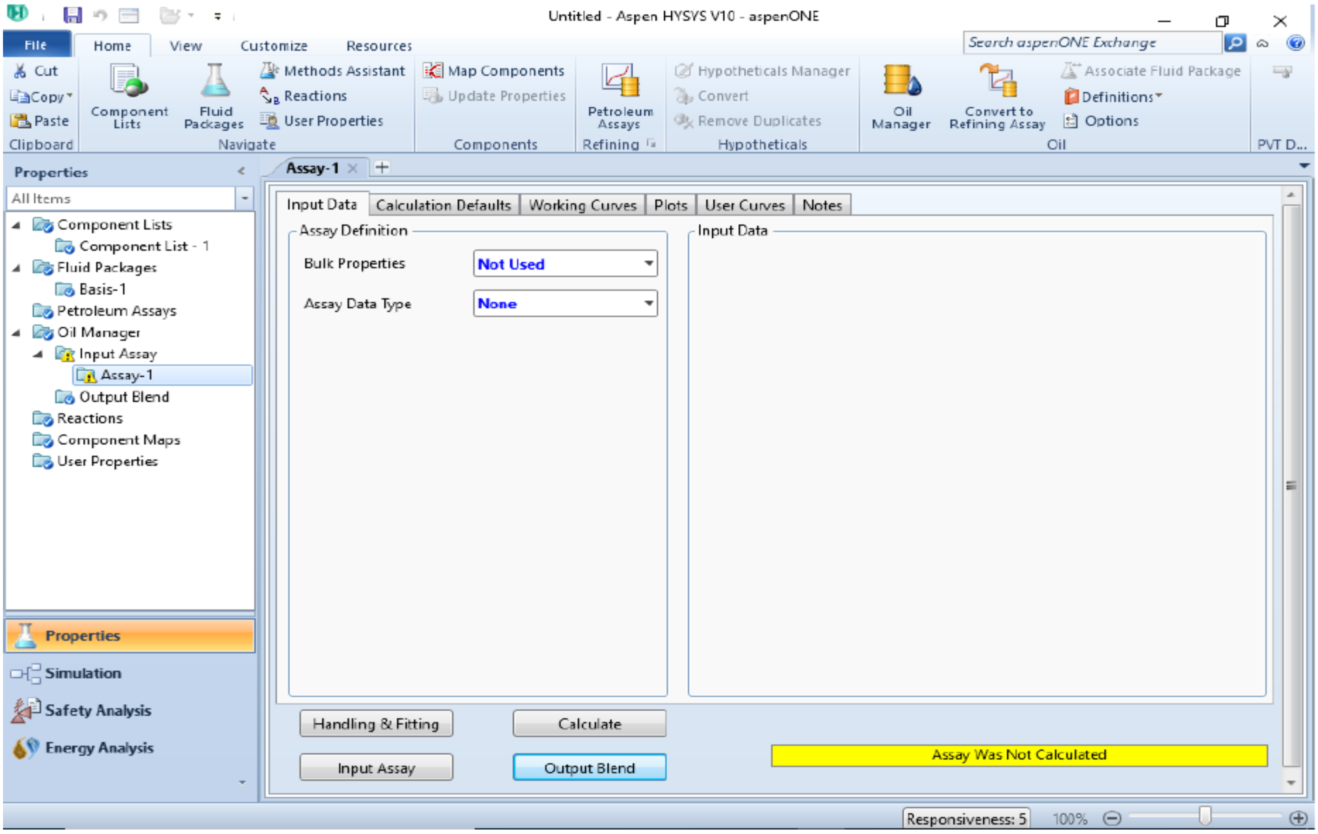


Figure 3.7: Oil manager selection



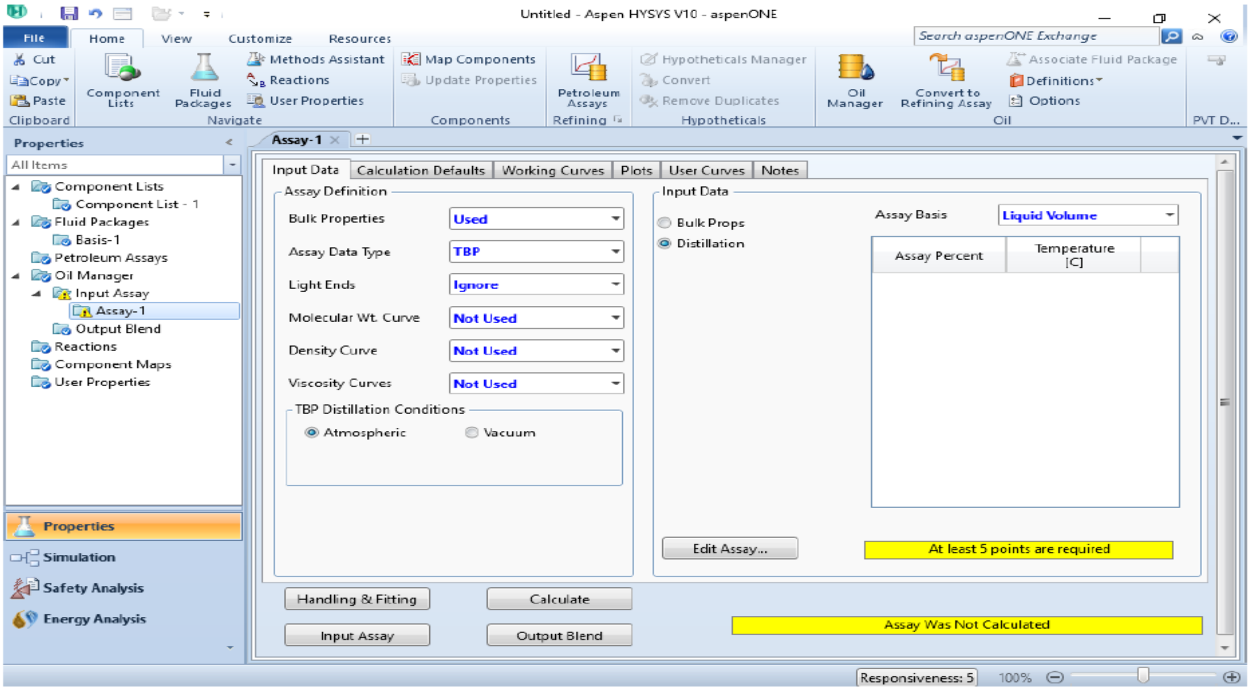
Bulk Properties: Not Used

Assay Data Type: None

Assay Was Not Calculated

Figure 3.8: Input assay from oil manager

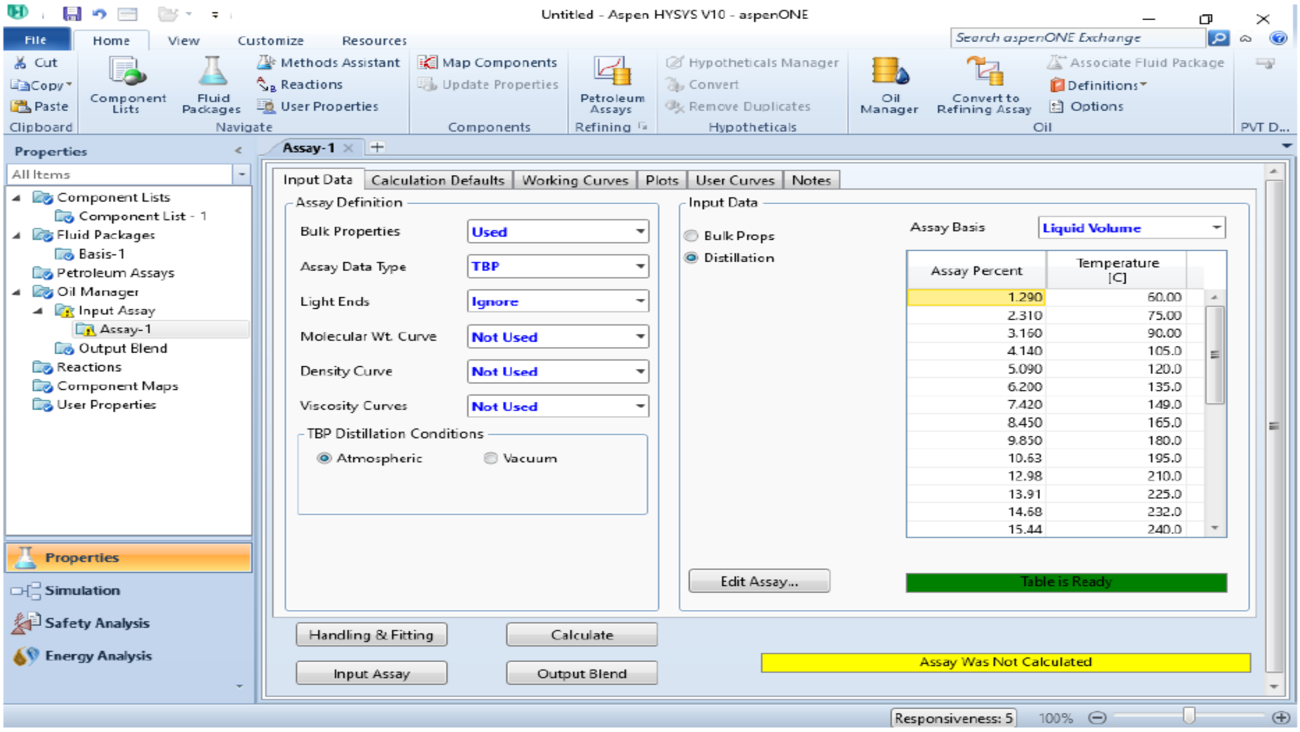
From input assay select bulk properties and activate to used, assay data type selects TBP and assay basis liquid volume (Figure 3.9).



Bulk Properties: Used
Assay Data Type: TBP
Light Ends: Ignore
Molecular Wt. Curve: Not Used
Density Curve: Not Used
Viscosity Curves: Not Used
TBP Distillation Conditions: ● Atmospheric ○ Vacuum
○ Bulk Props ● Distillation
Assay Basis: Liquid Volume
At least 5 points are required
Assay Was Not Calculated

Figure 3.9: Bulk properties and assay data type and assay basis

Then bulk properties data was chosen After that the, distillation curve data (TBP), were inserted (Figure 3.10),



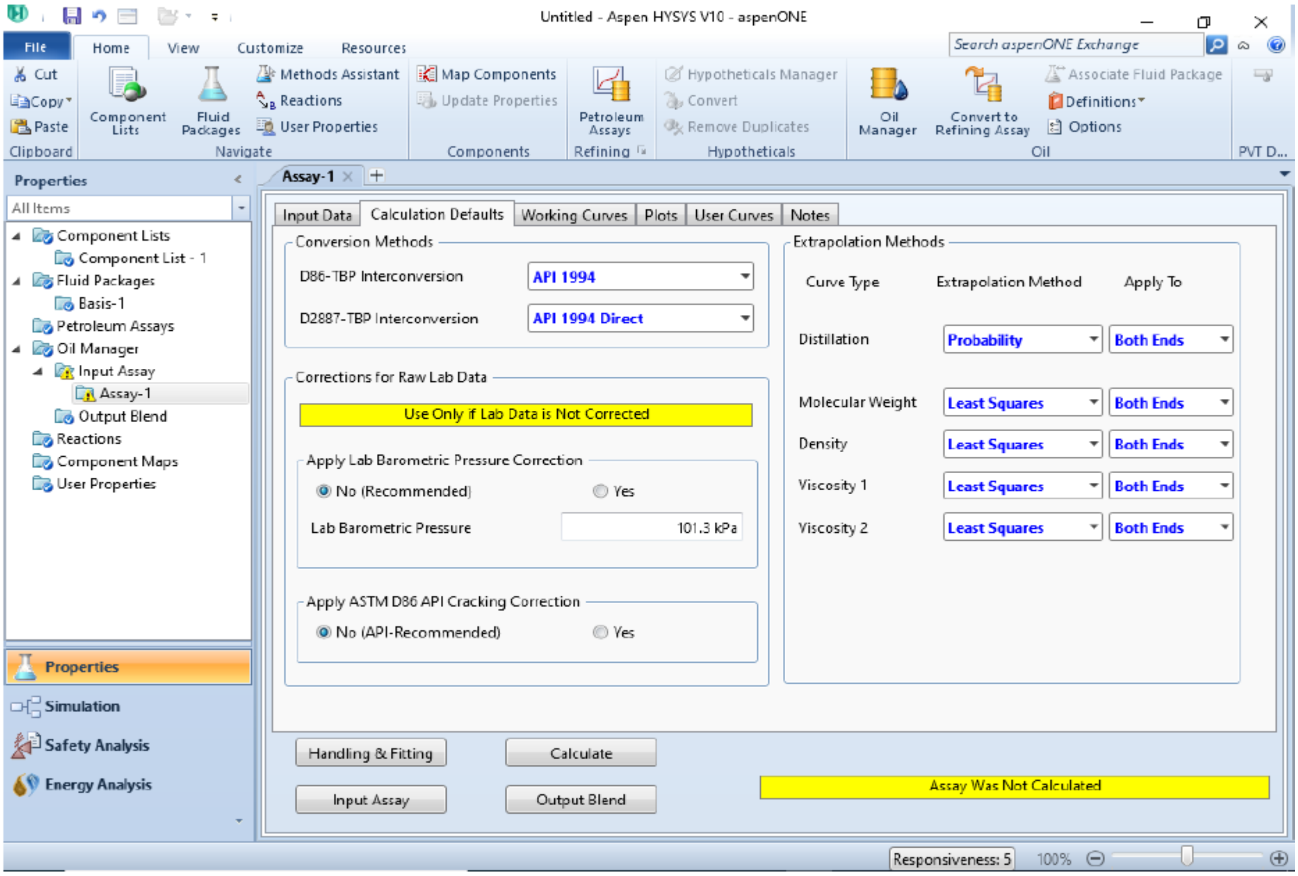
Bulk Properties: Used
Assay Data Type: TBP
Light Ends: Ignore
Molecular Wt. Curve: Not Used
Density Curve: Not Used
Viscosity Curves: Not Used
TBP Distillation Conditions: ● Atmospheric ○ Vacuum
○ Bulk Props ● Distillation
Assay Basis: Liquid Volume

Assay Percent	Temperature [C]
1.290	60.00
2.310	75.00
3.160	90.00
4.140	105.0
5.090	120.0
6.200	135.0
7.420	149.0
8.450	165.0
9.850	180.0
10.63	195.0
12.98	210.0
13.91	225.0
14.68	232.0
15.44	240.0

Table is Ready
Assay Was Not Calculated

Figure 3.10: Distillation curve data from TBP were inserted

Select calculation default in the extrapolation methods group, select probability for distillation curve type (Figure 3.11) and press calculate button at the bottom of the assay property view. HYSYS calculate the assay, and the status message at the bottom of the property view changes to assay was calculated (Figure 3.12).



Conversion Methods

D86-TBP Interconversion: API 1994
D2887-TBP Interconversion: API 1994 Direct

Corrections for Raw Lab Data: Use Only if Lab Data is Not Corrected

Apply Lab Barometric Pressure Correction: ● No (Recommended) ○ Yes
Lab Barometric Pressure: 101.3 kPa

Apply ASTM D86 API Cracking Correction: ● No (API-Recommended) ○ Yes

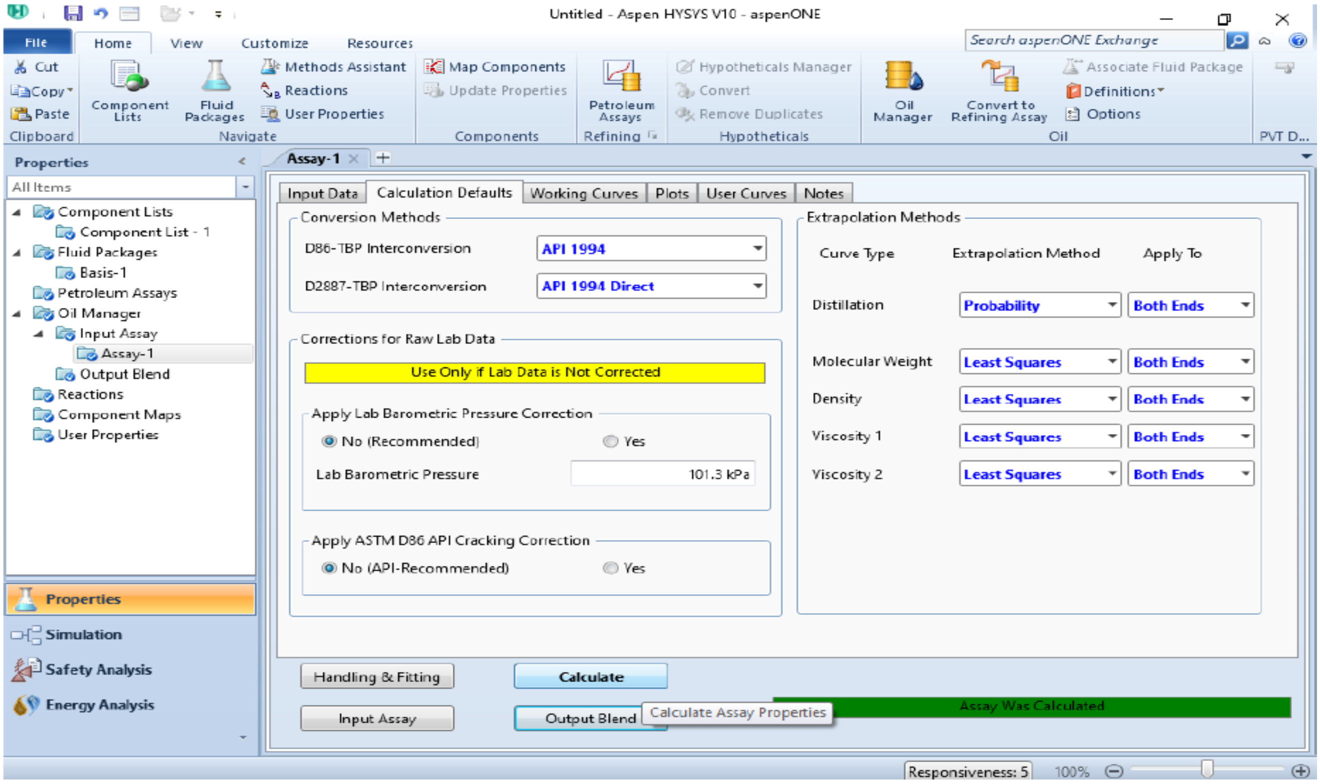
Extrapolation Methods

Curve Type	Extrapolation Method	Apply To
Distillation	Probability	Both Ends
Molecular Weight	Least Squares	Both Ends
Density	Least Squares	Both Ends
Viscosity 1	Least Squares	Both Ends
Viscosity 2	Least Squares	Both Ends

Assay Was Not Calculated

Figure 3.11: Probability selection

Now that the assay has been calculated, the next task is to cut the assay into individual petroleum hypocomponents (Figure 3.17). when the output blend is chosen click to add and transfer this assay for oil flow information (Figure3.14) and insert the flowrate to calculate new component (hypothetical components).



Conversion Methods

D86-TBP Interconversion: API 1994

D2887-TBP Interconversion: API 1994 Direct

Corrections for Raw Lab Data: Use Only if Lab Data is Not Corrected

Apply Lab Barometric Pressure Correction: ● No (Recommended) ○ Yes

Lab Barometric Pressure: 101.3 kPa

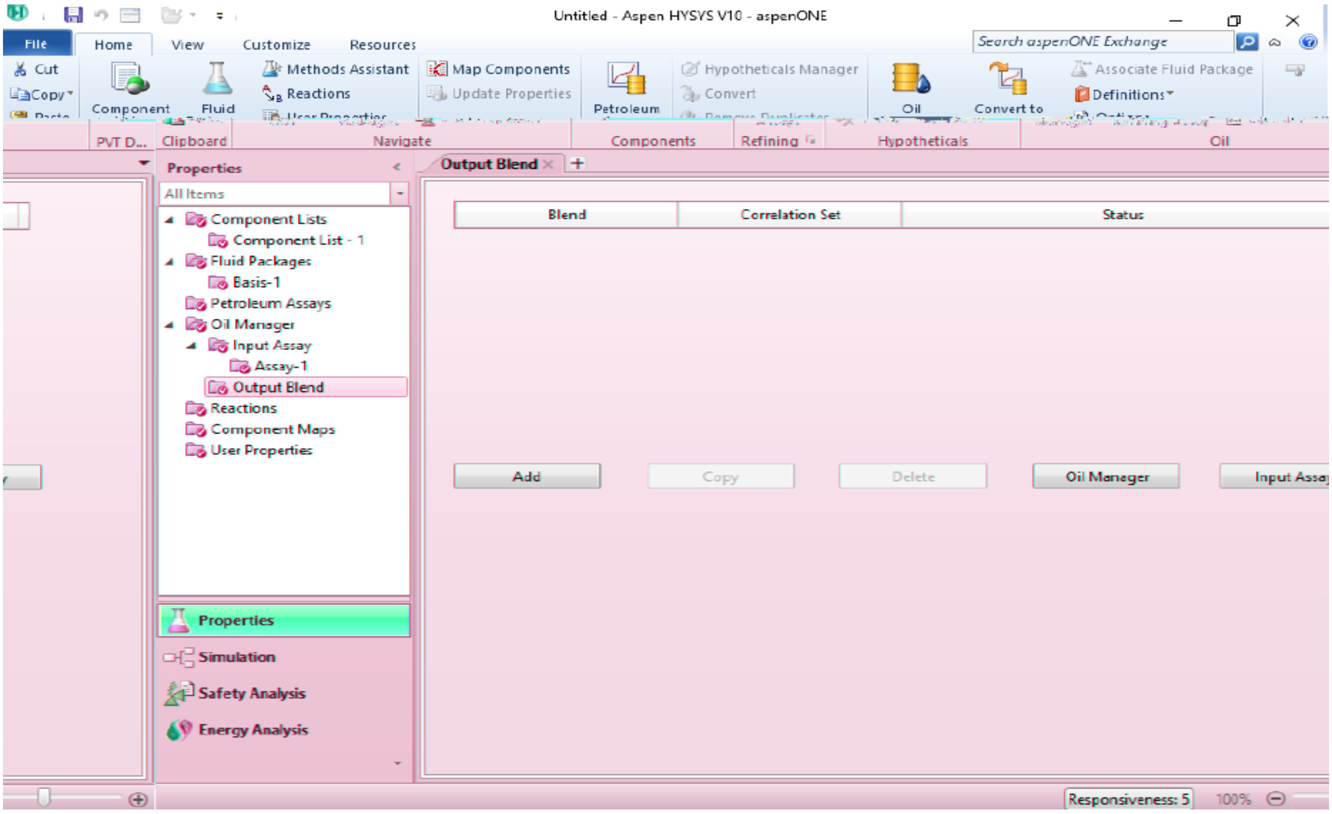
Apply ASTM D86 API Cracking Correction: ● No (API-Recommended) ○ Yes

Extrapolation Methods

Curve Type	Extrapolation Method	Apply To
Distillation	Probability	Both Ends
Molecular Weight	Least Squares	Both Ends
Density	Least Squares	Both Ends
Viscosity 1	Least Squares	Both Ends
Viscosity 2	Least Squares	Both Ends

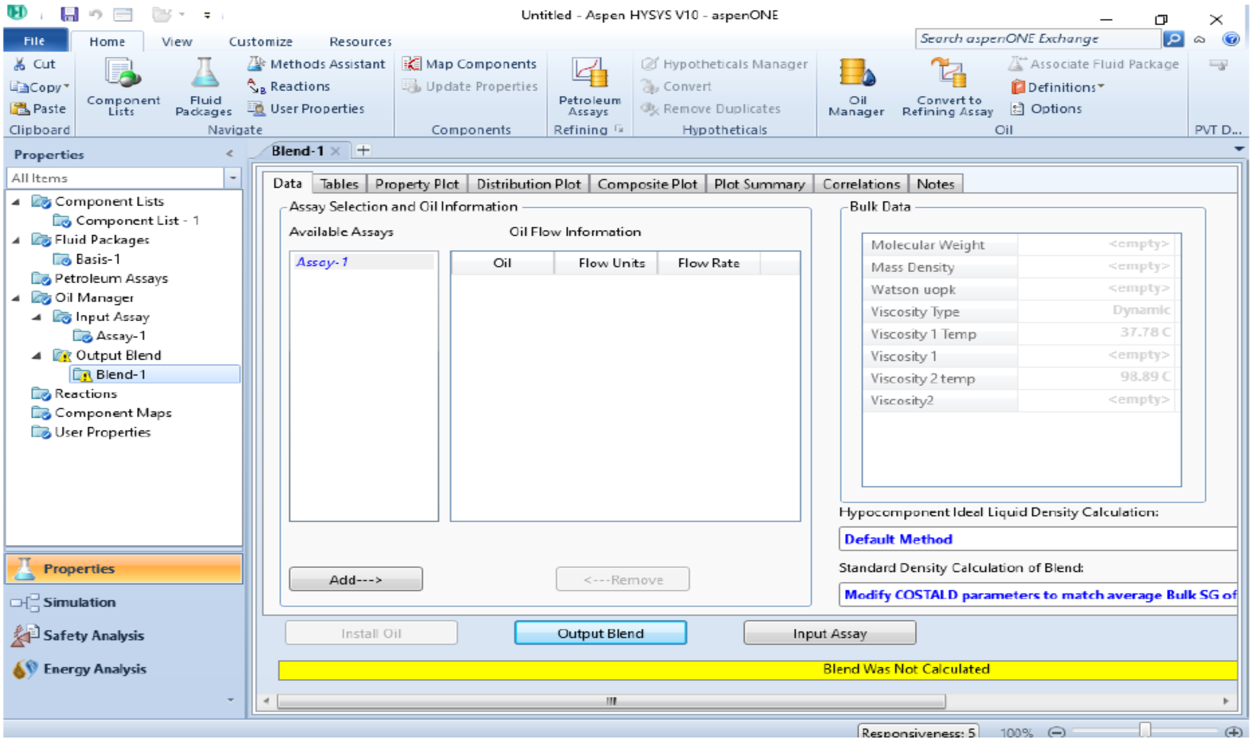
Assay Was Calculated

Figure 3.12: Assay was calculated



Blend	Correlation Set	Status

Figure 3.13: Create the output blend



Oil	Flow Units	Flow Rate

Bulk Data

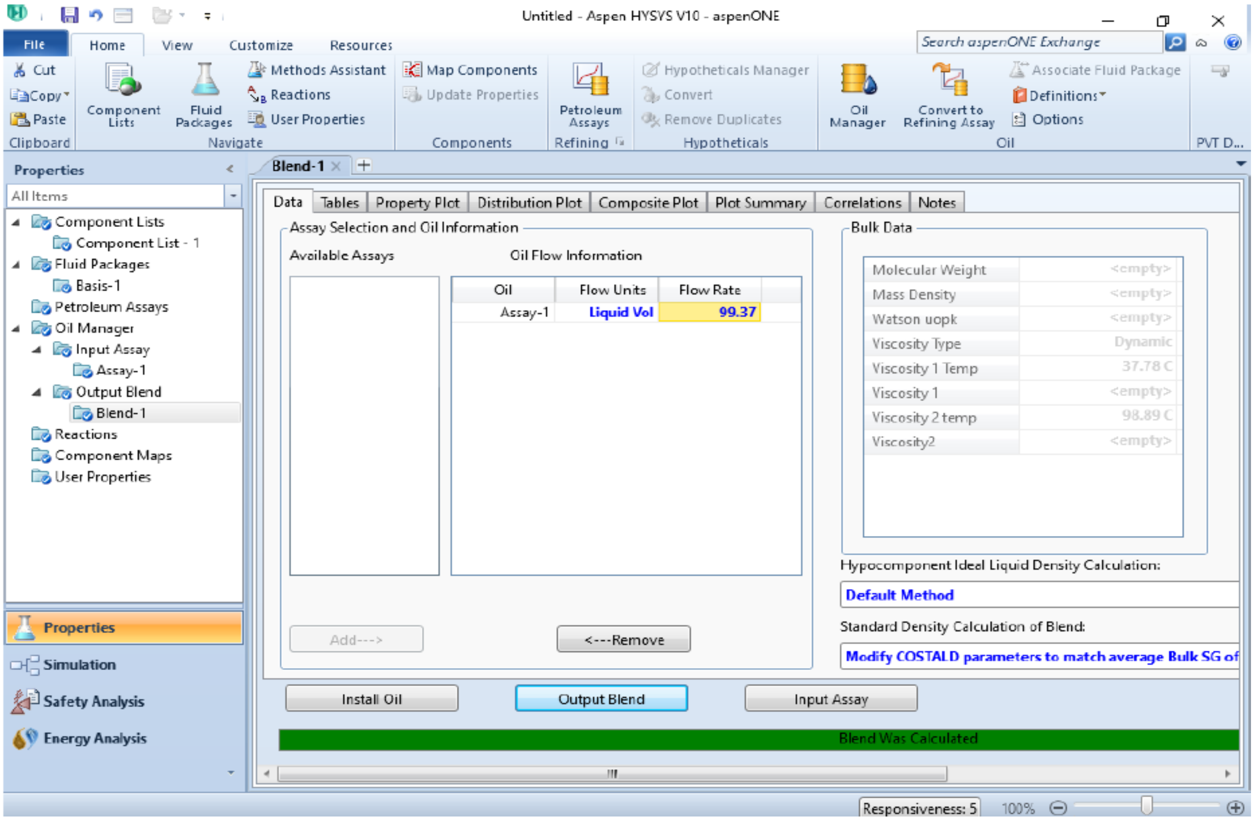
Molecular Weight	<empty>
Mass Density	<empty>
Watson uopk	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	37.78 C
Viscosity 1	<empty>
Viscosity 2 temp	98.89 C
Viscosity2	<empty>

Hypocomponent Ideal Liquid Density Calculation: Default Method

Standard Density Calculation of Blend: Modify COSTALD parameters to match average Bulk SG of

Blend Was Not Calculated

Figure 3.14: Assay transfer for oil flow information



Oil	Flow Units	Flow Rate
Assay-1	Liquid Vol	99.37

Bulk Data

Molecular Weight	<empty>
Mass Density	<empty>
Watson uopk	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	37.78 C
Viscosity 1	<empty>
Viscosity 2 temp	98.89 C
Viscosity2	<empty>

Hypocomponent Ideal Liquid Density Calculation: Default Method

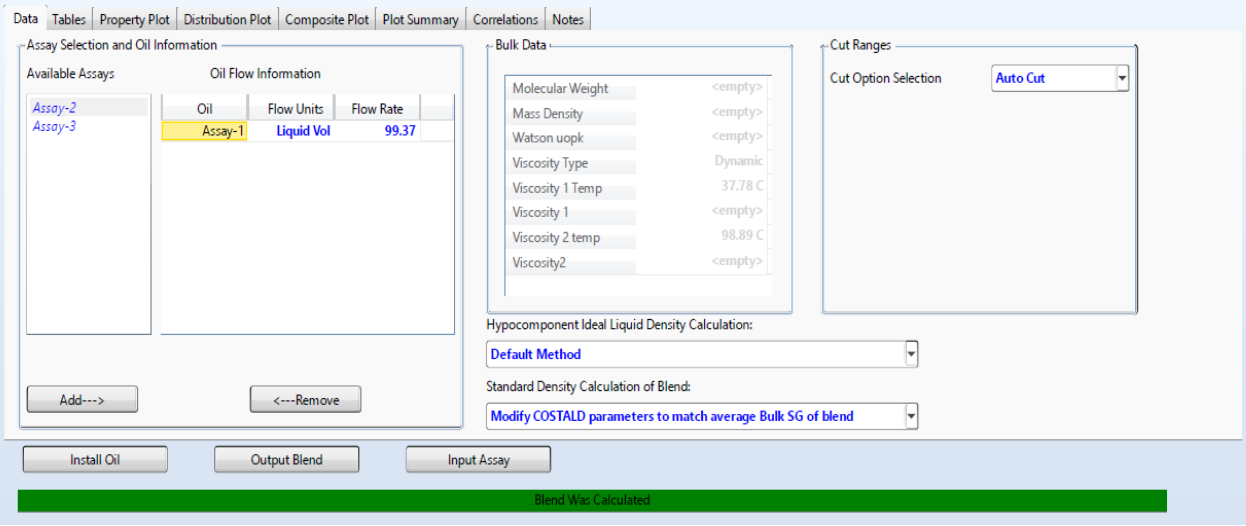
Standard Density Calculation of Blend: Modify COSTALD parameters to match average Bulk SG of

Blend Was Calculated

Figure 3.15: Assay was calculated

A blend (Cut) is automatically calculated based on the current cut option.

auto blend option was used for cut option(Figure3.16)



Available Assays: Assay-2, Assay-3

Oil	Flow Units	Flow Rate
Assay-1	Liquid Vol	99.37

Bulk Data	
Molecular Weight	<empty>
Mass Density	<empty>
Watson uopk	<empty>
Viscosity Type	Dynamic
Viscosity 1 Temp	37.78 C
Viscosity 1	<empty>
Viscosity 2 temp	98.89 C
Viscosity2	<empty>

Cut Option Selection: Auto Cut

Hypocomponent Ideal Liquid Density Calculation: Default Method

Standard Density Calculation of Blend: Modify COSTALD parameters to match average Bulk SG of blend

Blend Was Calculated

Figure 3.16: Cut option selection

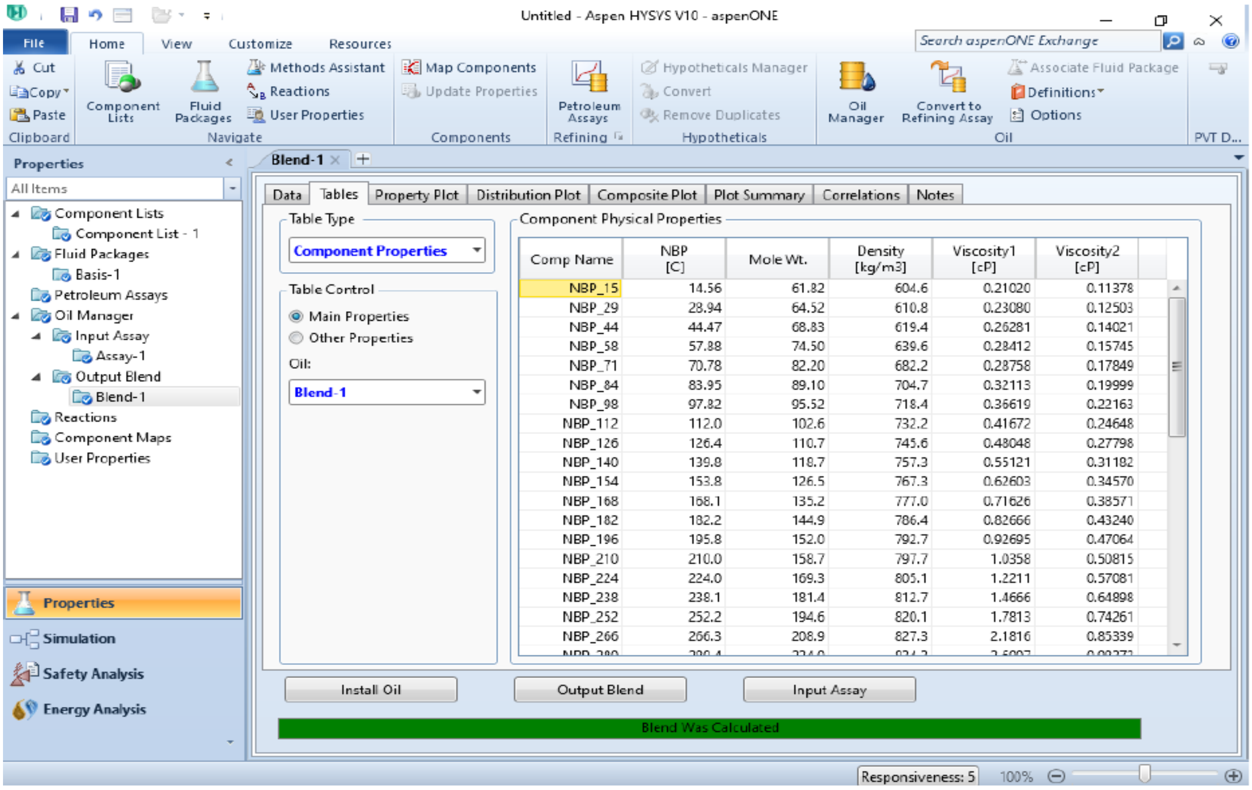


Table Type: Component Properties

Table Control: ● Main Properties ○ Other Properties

Oil: Blend-1

Comp Name	NBP [C]	Mole Wt.	Density [kg/m3]	Viscosity1 [cP]	Viscosity2 [cP]
NBP_15	14.56	61.82	604.6	0.21020	0.11378
NBP_29	28.94	64.52	610.8	0.23080	0.12503
NBP_44	44.47	68.83	619.4	0.26281	0.14021
NBP_58	57.88	74.50	639.6	0.28412	0.15745
NBP_71	70.78	82.20	682.2	0.28758	0.17849
NBP_84	83.95	89.10	704.7	0.32113	0.19999
NBP_98	97.82	95.52	718.4	0.36619	0.22163
NBP_112	112.0	102.6	732.2	0.41672	0.24648
NBP_126	126.4	110.7	745.6	0.48048	0.27798
NBP_140	139.8	118.7	757.3	0.55121	0.31182
NBP_154	153.8	126.5	767.3	0.62603	0.34570
NBP_168	168.1	135.2	777.0	0.71626	0.38571
NBP_182	182.2	144.9	786.4	0.82666	0.43240
NBP_196	195.8	152.0	792.7	0.92695	0.47064
NBP_210	210.0	158.7	797.7	1.0358	0.50815
NBP_224	224.0	169.3	805.1	1.2211	0.57081
NBP_238	238.1	181.4	812.7	1.4666	0.64898
NBP_252	252.2	194.6	820.1	1.7813	0.74261
NBP_266	266.3	208.9	827.3	2.1816	0.85339

Blend Was Calculated

Figure 3.17: Main properties of hypothetical components

At the bottom of the cut input information group, the user custom radio button is selected, and HYSYS provides default TBP cut point temperatures for each user custom product. The cut distributions table shows the fraction of each product in the blend. Since liquid volume is the current basis in the table control group, the products are listed according to liquid volume fraction (figure3.18). These fractions can be used to estimate the products flow rate for the fractionation column.

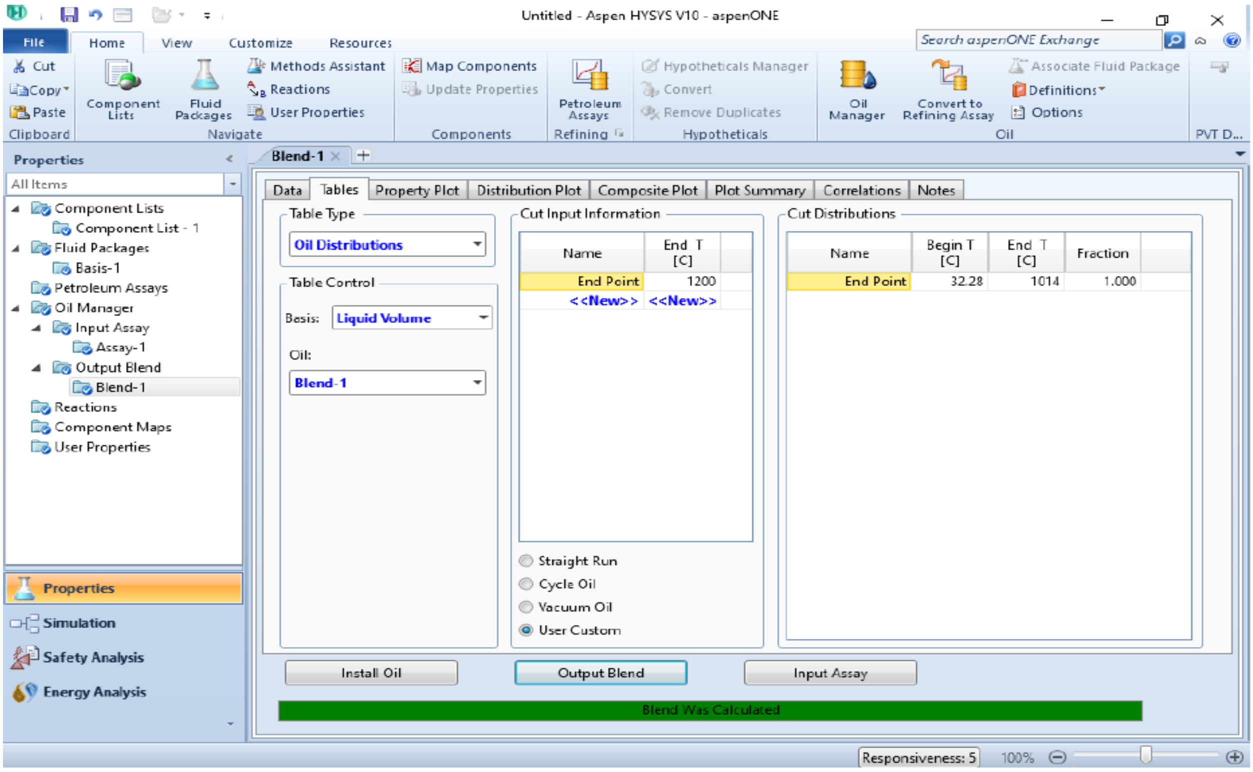


Table Type: Oil Distributions

Table Control

Basis: Liquid Volume

Oil: Blend-1

Cut Input Information

Name	End T [C]
End Point	1200
<<New>>	<<New>>

○ Straight Run
○ Cycle Oil
○ Vacuum Oil
● User Custom

Cut Distributions

Name	Begin T [C]	End T [C]	Fraction
End Point	32.28	1014	1.000

Blend Was Calculated

Figure 3.18: Cut distribution for liquid volume

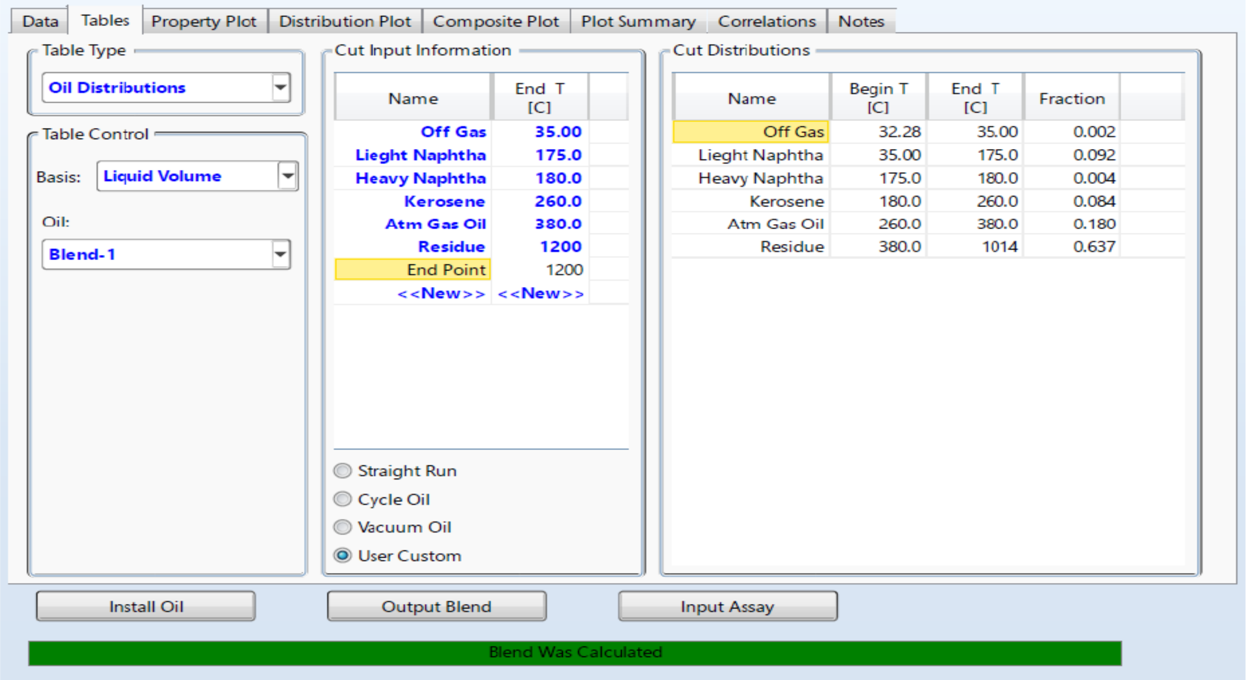


Table Type: Oil Distributions

Table Control

Basis: Liquid Volume

Oil: Blend-1

Cut Input Information

Name	End T [C]
Off Gas	35.00
Lieght Naphtha	175.0
Heavy Naphtha	180.0
Kerosene	260.0
Atm Gas Oil	380.0
Residue	1200
End Point	1200
<<New>>	<<New>>

○ Straight Run
○ Cycle Oil
○ Vacuum Oil
● User Custom

Cut Distributions

Name	Begin T [C]	End T [C]	Fraction
Off Gas	32.28	35.00	0.002
Lieght Naphtha	35.00	175.0	0.092
Heavy Naphtha	175.0	180.0	0.004
Kerosene	180.0	260.0	0.084
Atm Gas Oil	260.0	380.0	0.180
Residue	380.0	1014	0.637

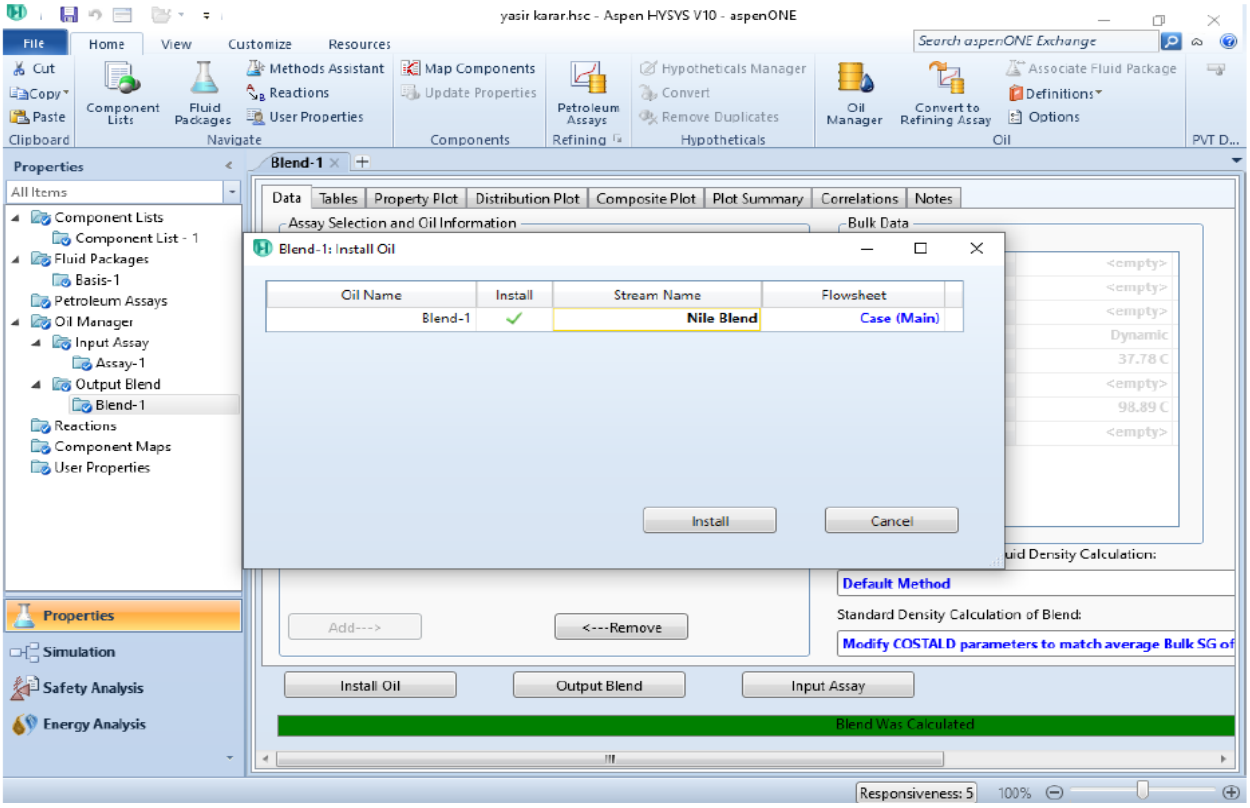
Blend Was Calculated

Figure 3.19: Liquid volume fraction of products from oil cut distribution

The last step in the oil characterization procedure is to install the oil and rename, which accomplishes the following (Figure3.20):

- The petroleum hypocomponents are added to the fluid package.
- The calculated light ends and oil composition are transferred to a material stream for use in the simulation.

By applying the same method, all the properties and proportion of Rawat and Thargath crude were calculated, and 7mixtures of the previous crude were made in different proportion.



Oil Name	Install	Stream Name	Flowsheet
Blend-1	✓	Nile Blend	Case (Main)

<empty>
<empty>
<empty>
Dynamic
37.78 C
<empty>
98.89 C
<empty>

uid Density Calculation: Default Method

Standard Density Calculation of Blend: Modify COSTALD parameters to match average Bulk SG of

Blend Was Calculated

Figure 3.20: Install the oil and rename

In general, the first task in the simulation environment is to install one or more feed streams, however, the stream preheat crude was already installed during the oil characterization procedure (Figure3.21). At this point, the compositions tab of the workbook property view was very important.

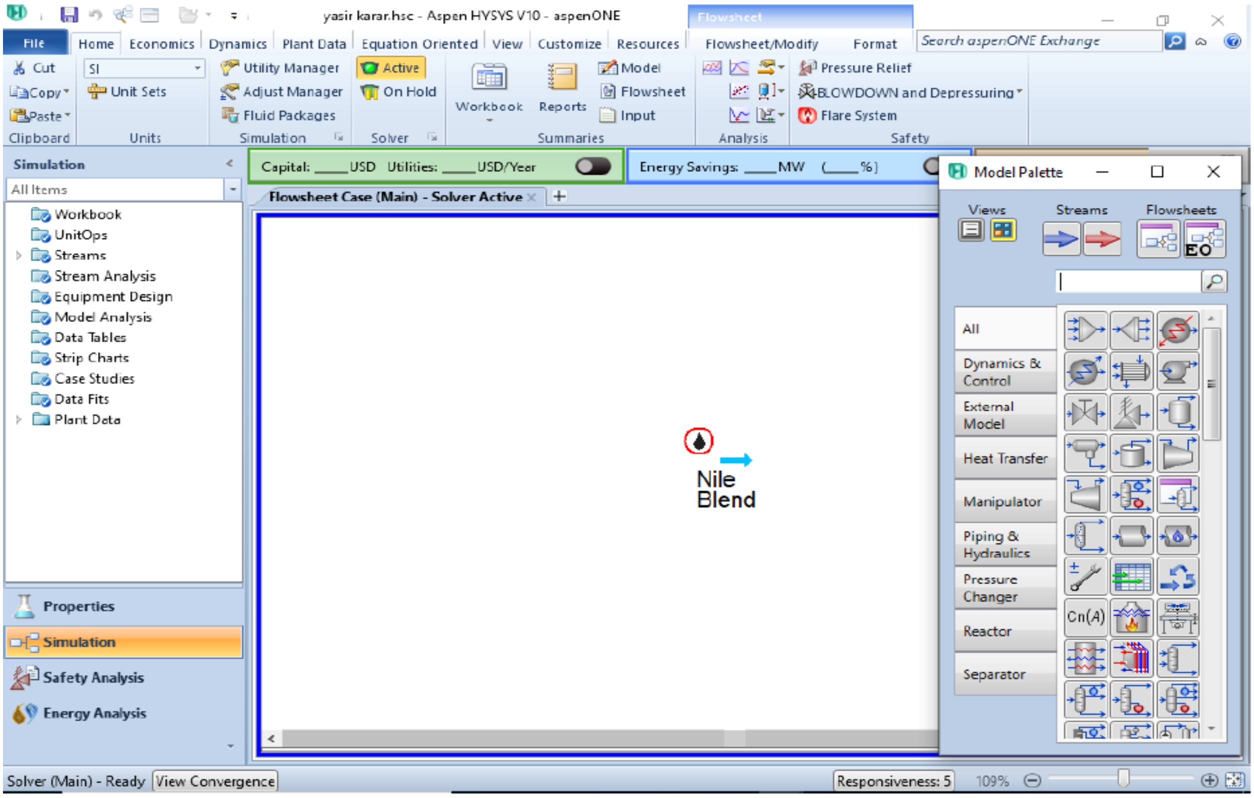
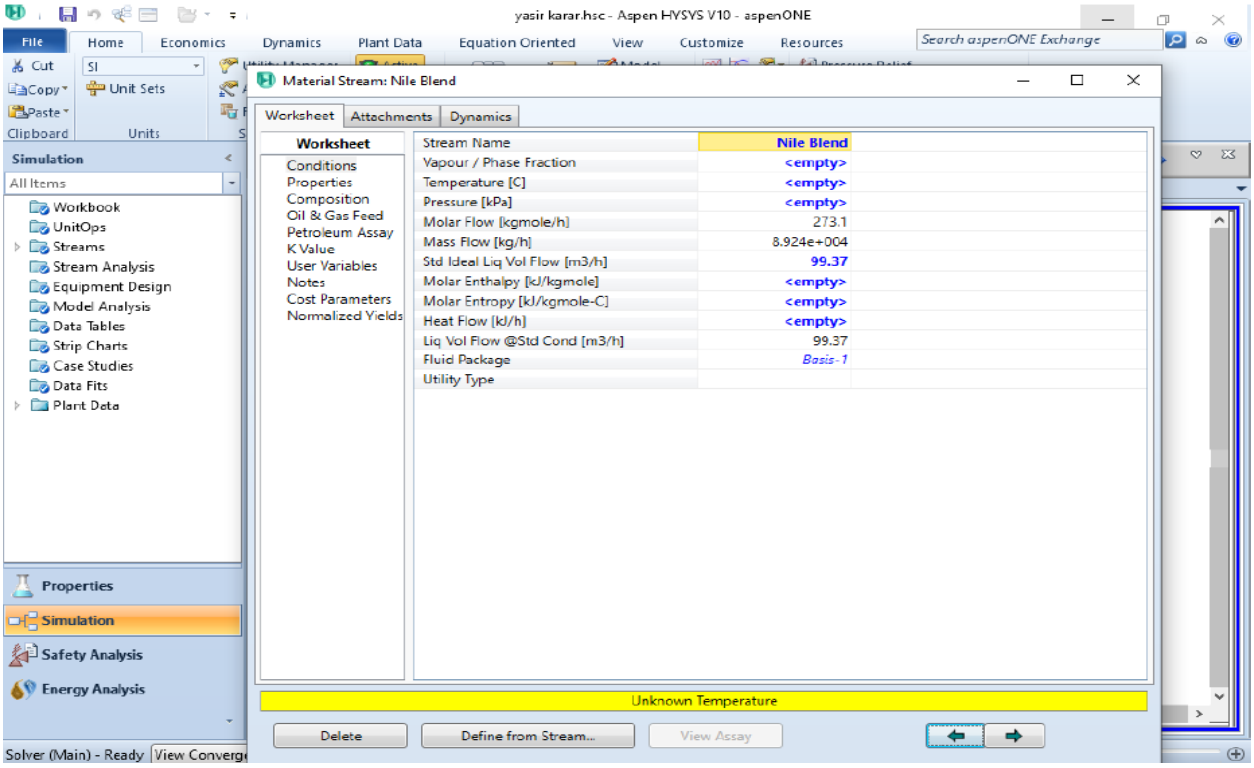


Figure 3.21: Liquid volume fraction of products from oil distribution

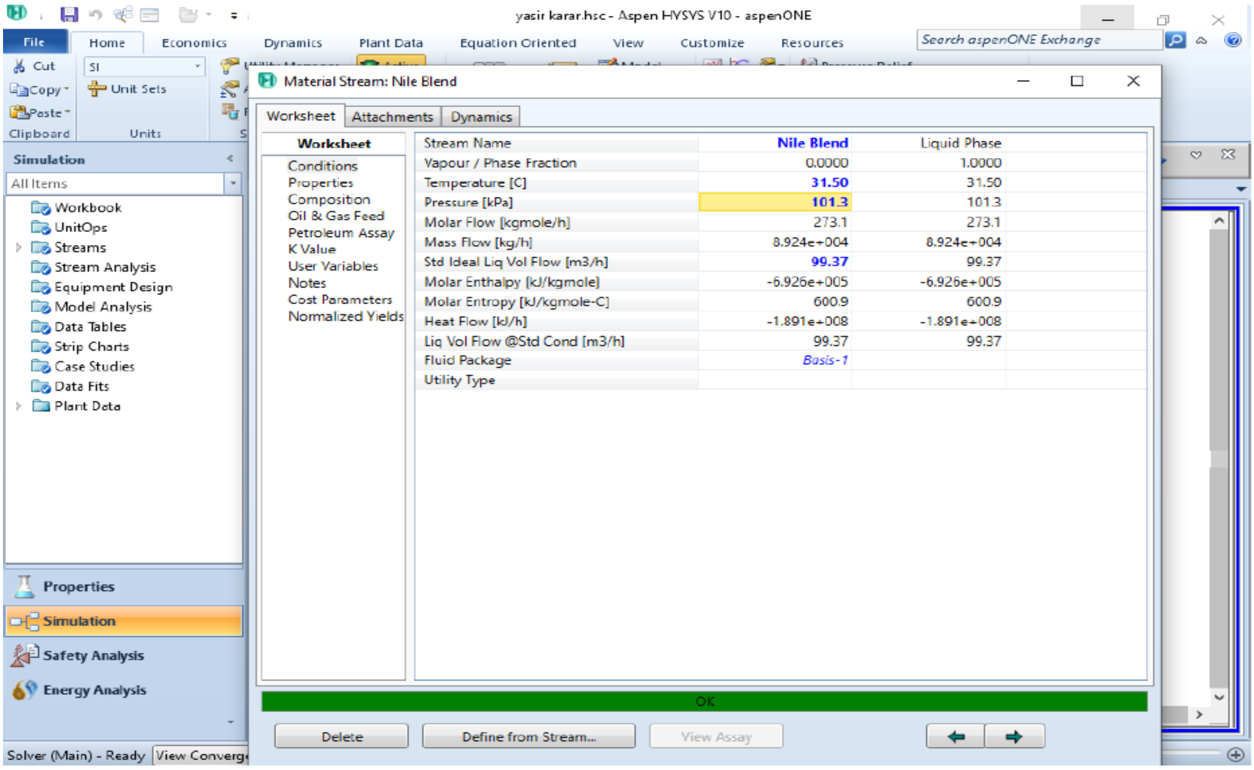
click on the arrow (Nile Blend), define this stream by adding the temperature and pressure (Figure3.22).



Stream Name	Nile Blend
Vapour / Phase Fraction	<empty>
Temperature [C]	<empty>
Pressure [kPa]	<empty>
Molar Flow [kgmole/h]	273.1
Mass Flow [kg/h]	8.924e+004
Std Ideal Liq Vol Flow [m3/h]	99.37
Molar Enthalpy [kJ/kgmole]	<empty>
Molar Entropy [kJ/kgmole-C]	<empty>
Heat Flow [kJ/h]	<empty>
Liq Vol Flow @Std Cond [m3/h]	99.37
Fluid Package	Basis-1
Utility Type	

Unknown Temperature

Figure 3.22: Definition of the Nile Blend by temperature and pressure



Stream Name	Nile Blend	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature [C]	31.50	31.50
Pressure [kPa]	101.3	101.3
Molar Flow [kgmole/h]	273.1	273.1
Mass Flow [kg/h]	8.924e+004	8.924e+004
Std Ideal Liq Vol Flow [m3/h]	99.37	99.37
Molar Enthalpy [kJ/kgmole]	-6.926e+005	-6.926e+005
Molar Entropy [kJ/kgmole-C]	600.9	600.9
Heat Flow [kJ/h]	-1.891e+008	-1.891e+008
Liq Vol Flow @Std Cond [m3/h]	99.37	99.37
Fluid Package	Basis-1	
Utility Type		

Figure 3.23: Nile Blend were definition

The preheated crude enters the fractionation column at 371.1°C and 482.6 kPa. for this reason, the first step is to create a new heater on the flow sheet for heating the crude oil because the normal condition for crude temperature 31.5°C and pressure 101.3 kPa. In the model palette, click the heater icon. create a new heater(Figure3.24). Double click the Heater icon to open its property view. Click the design tab, then select the connections page. The names of the inlet (Nile Blend), outlet (atmospheric Feed), and energy streams (Heater Duty) (Figure3.25).

insert temperature 371.1°C and pressure 482.6 kPa, the required heating duty calculated by HYSYS appears in the heat flow cell(Figure3.26).

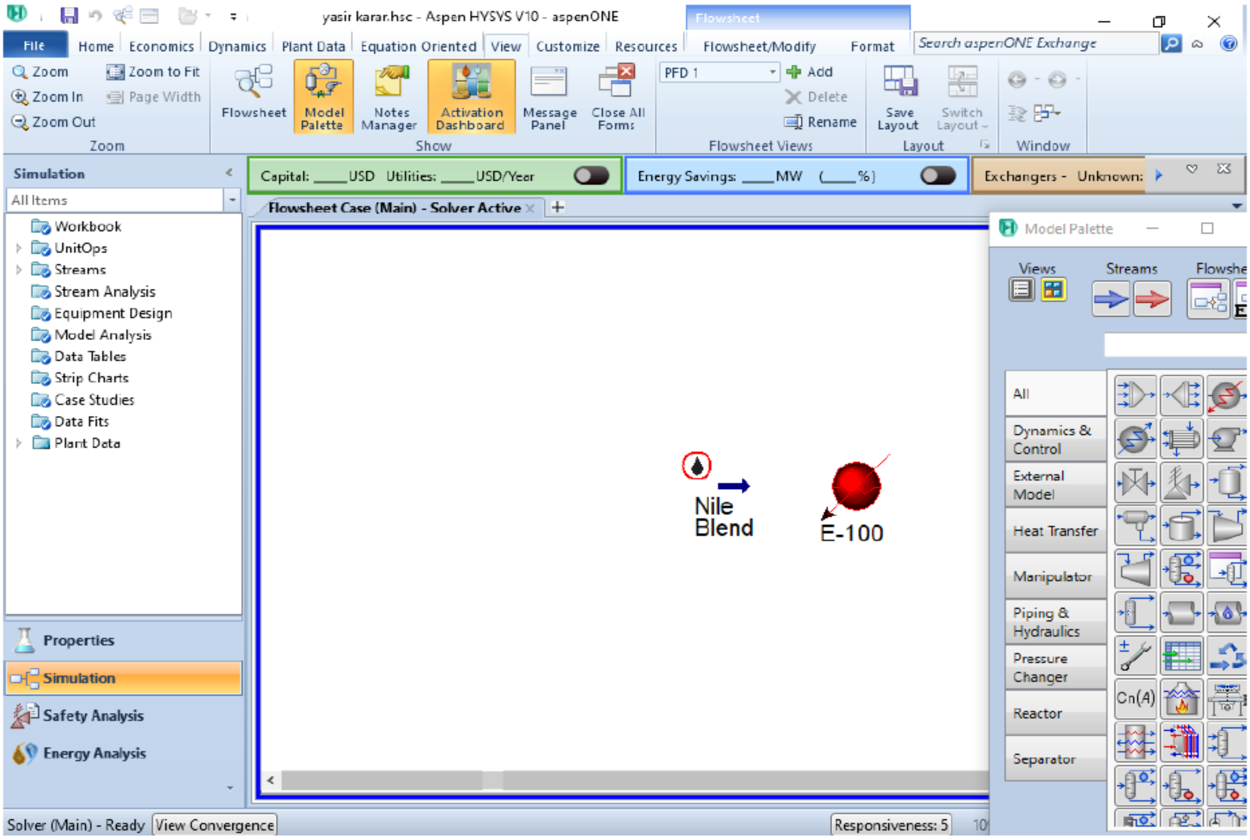
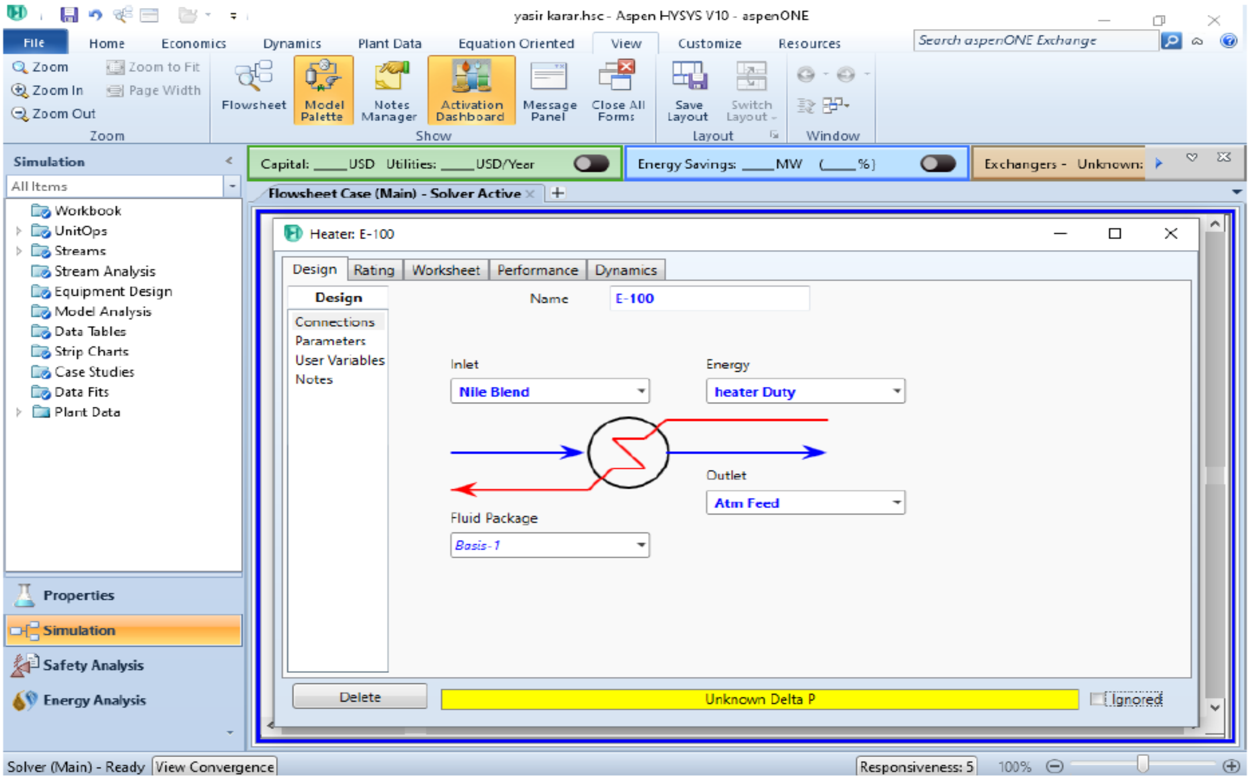


Figure 3.24: Heater addition



Name: E-100

Inlet: Nile Blend

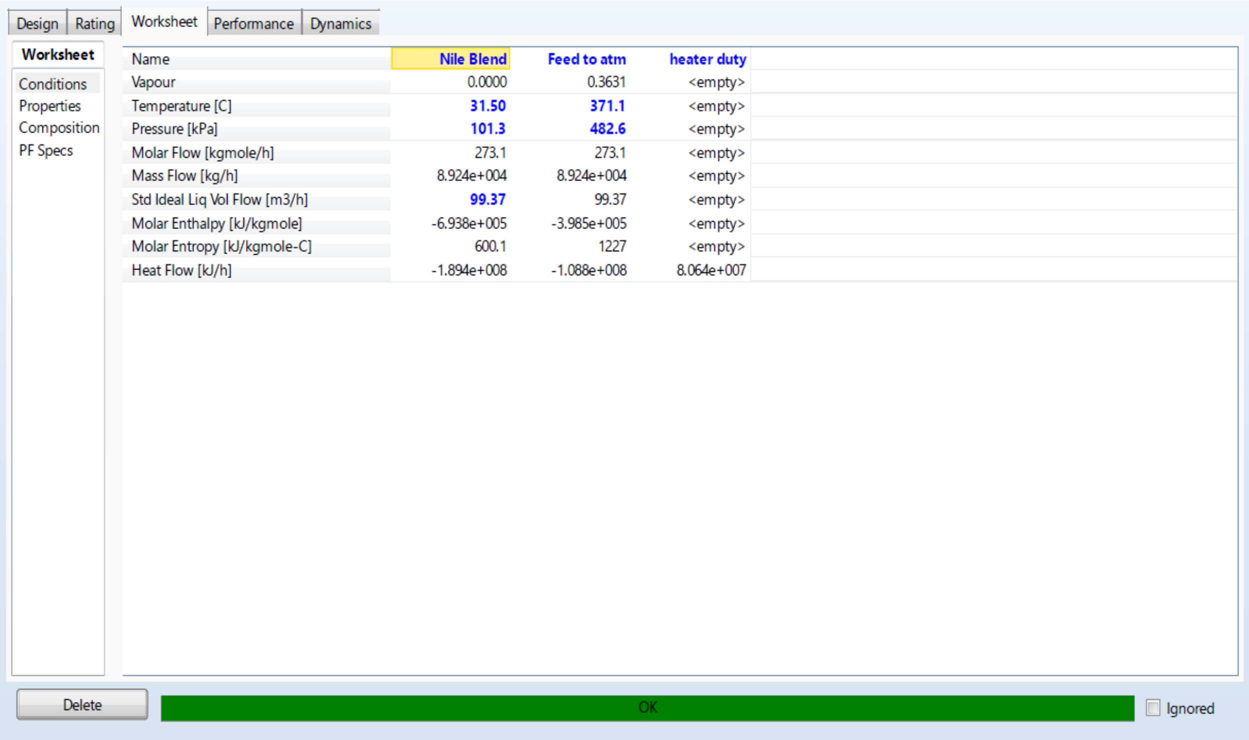
Energy: heater Duty

Outlet: Atm Feed

Fluid Package: Basis-1

Unknown Delta P

Figure 3.25: Modifying heater properties



Name	Nile Blend	Feed to atm	heater duty
Vapour	0.0000	0.3631	<empty>
Temperature [C]	31.50	371.1	<empty>
Pressure [kPa]	101.3	482.6	<empty>
Molar Flow [kgmole/h]	273.1	273.1	<empty>
Mass Flow [kg/h]	8.924e+004	8.924e+004	<empty>
Std Ideal Liq Vol Flow [m3/h]	99.37	99.37	<empty>
Molar Enthalpy [kJ/kgmole]	-6.938e+005	-3.985e+005	<empty>
Molar Entropy [kJ/kgmole-C]	600.1	1227	<empty>
Heat Flow [kJ/h]	-1.894e+008	-1.088e+008	8.064e+007

Figure 3.26: The heater was installation

after the heater installation, create four steam streams (Bottom steam, Heavy Naphtha steam, Kerosene steam, Atmospheric Gas Oil steam) and energy stream (Trim Duty) (Figure 3.27), To add the first utility stream, double click the material stream icon on the model palette. In the stream property view, click in the stream name cell and rename the stream heavy naphtha steam (Figure 3.28). and define the conditions (Temperature, Pressure and Flowrate) of this stream (Figure 3.29), now that the utility stream conditions have been specified, the next task is to input the compositions. Click the compositions tab in the workbook. The components are listed by mole fraction by default (Figure 3.30), change the basis for liquid volume fraction (Figure 3.31). Click in the input cell for the first component H_2O, since the stream is all water, type 1 for the H_2O then press enter. The input composition for stream properties appears. By repeating these steps for the steam streams, their components were definition (Figure 3.32).

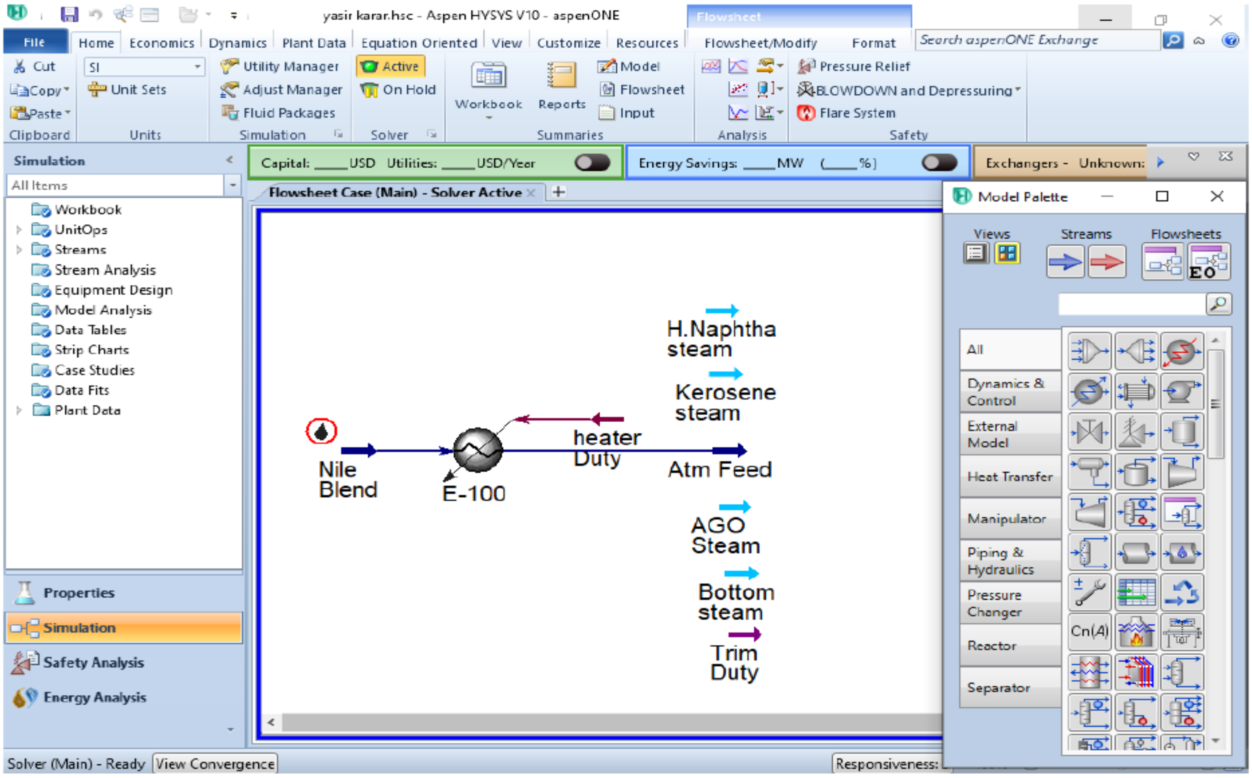
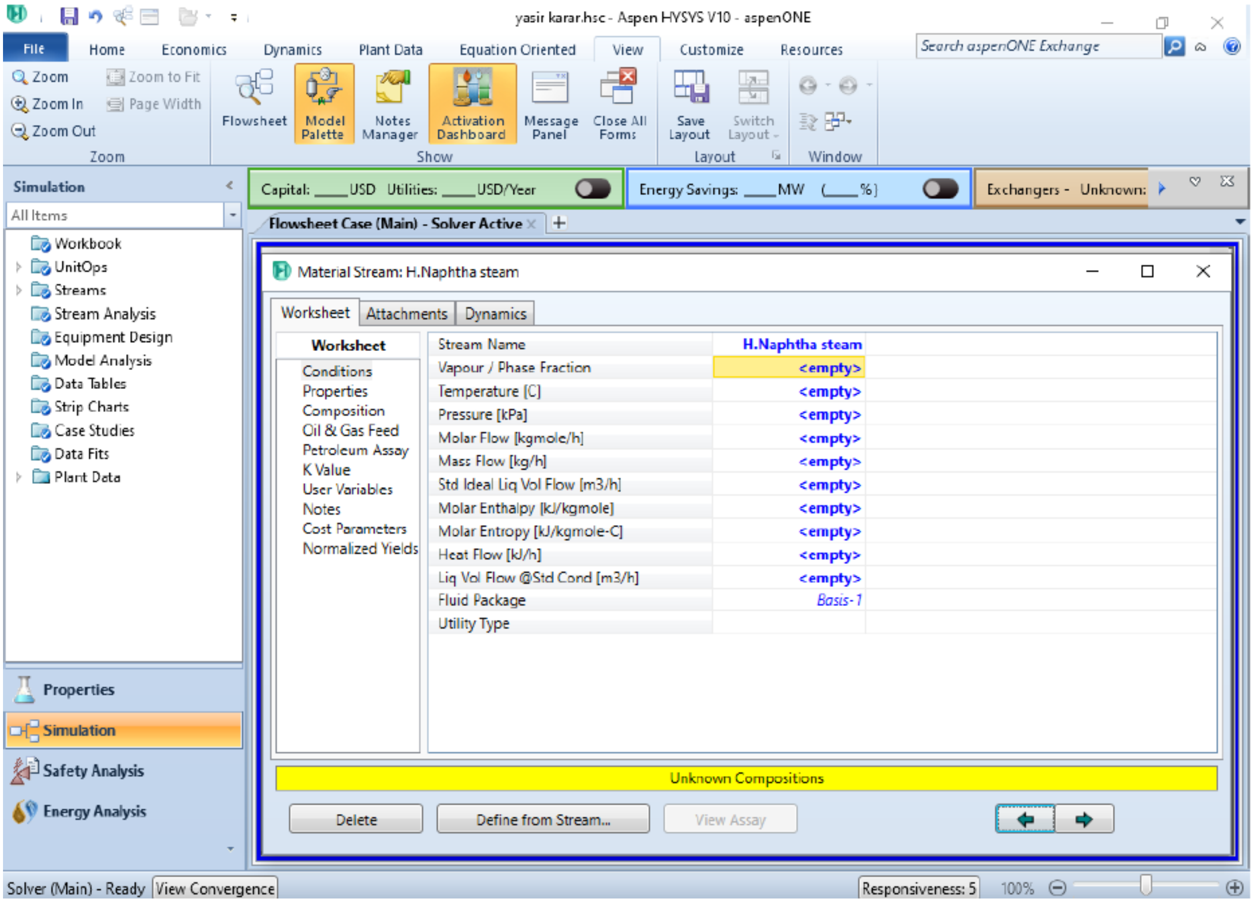


Figure 3.27: Addition four steam stream and trim duty

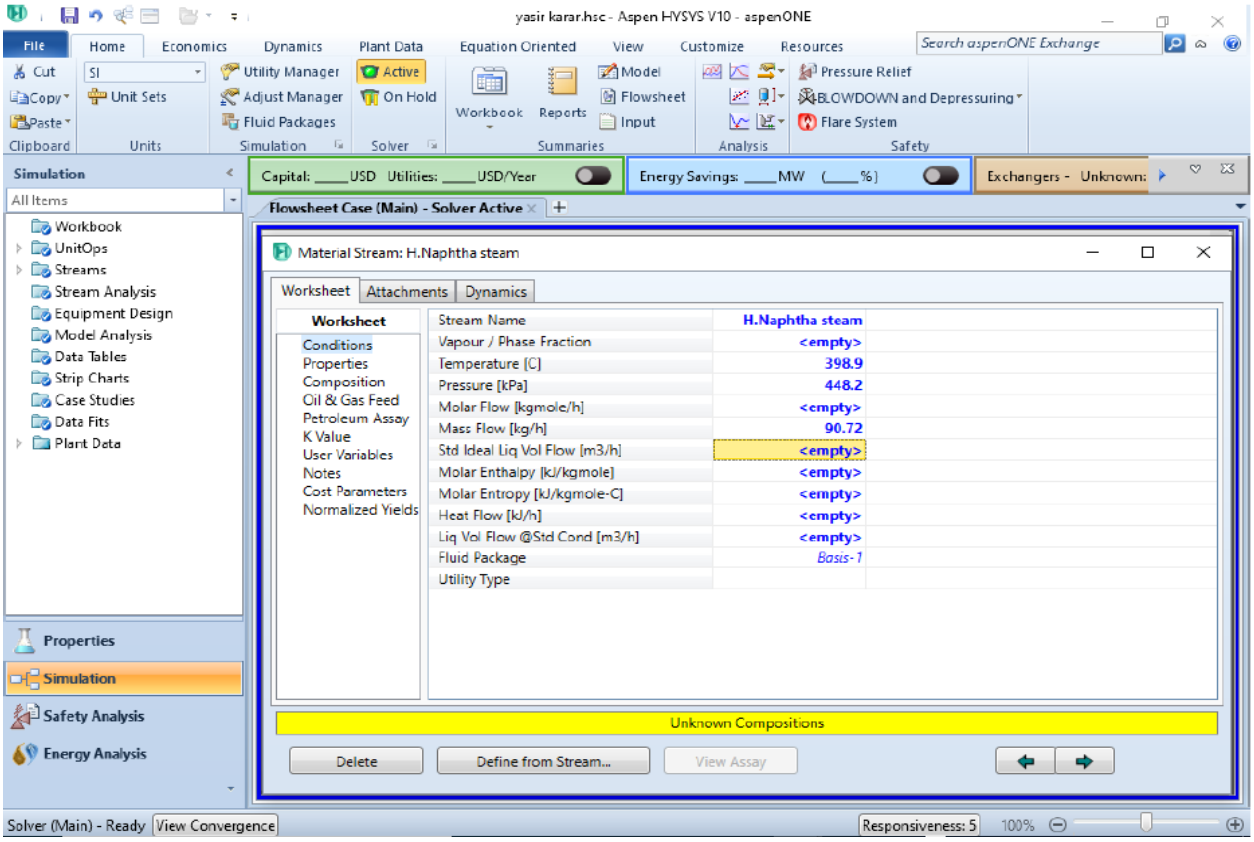


Material Stream: H.Naphtha steam

Stream Name	H.Naphtha steam
Vapour / Phase Fraction	<empty>
Temperature [C]	<empty>
Pressure [kPa]	<empty>
Molar Flow [kgmole/h]	<empty>
Mass Flow [kg/h]	<empty>
Std Ideal Liq Vol Flow [m3/h]	<empty>
Molar Enthalpy [kJ/kgmole]	<empty>
Molar Entropy [kJ/kgmole-C]	<empty>
Heat Flow [kJ/h]	<empty>
Liq Vol Flow @Std Cond [m3/h]	<empty>
Fluid Package	Basis-1
Utility Type	

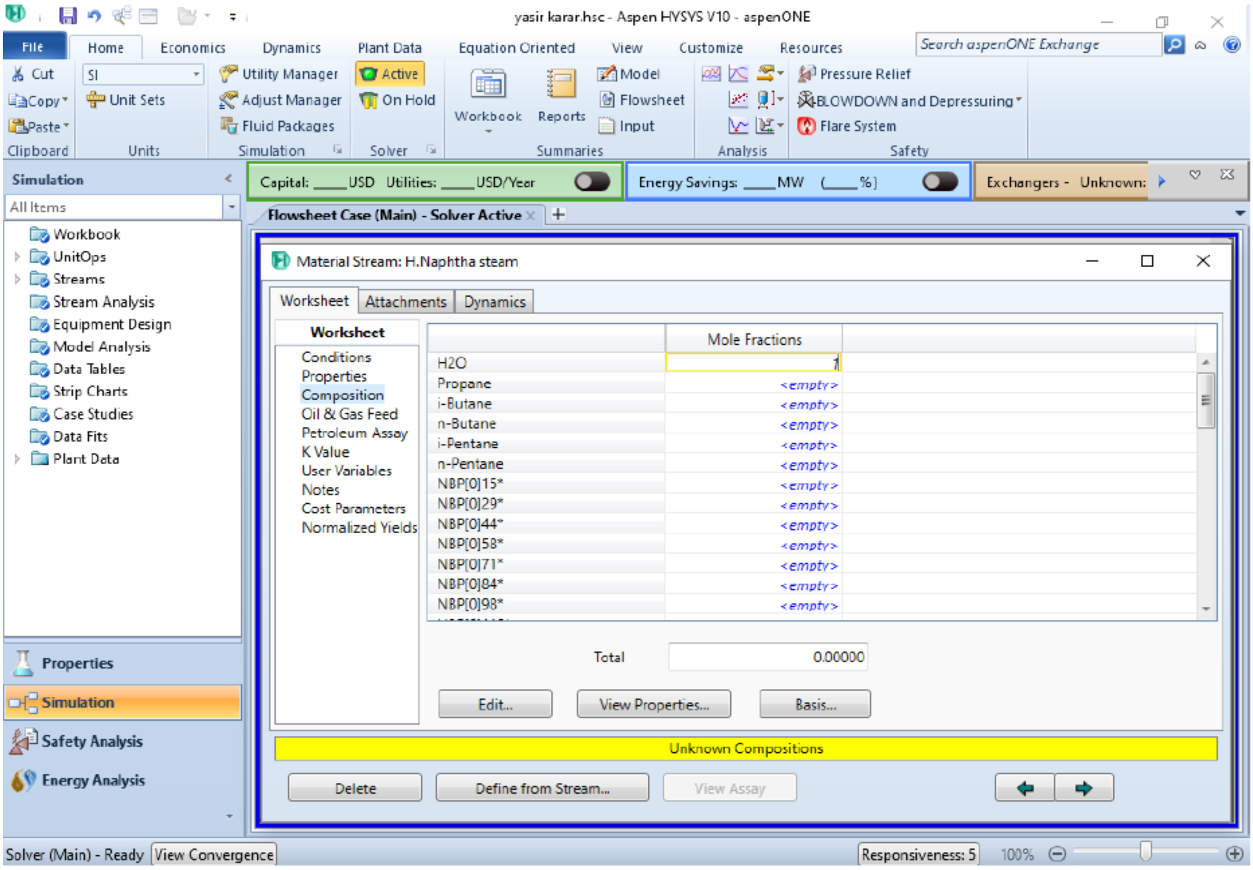
Unknown Compositions

Figure 3.28: Heavy Naphtha steam stream from flowsheet



Stream Name	H.Naphtha steam
Vapour / Phase Fraction	<empty>
Temperature [C]	398.9
Pressure [kPa]	448.2
Molar Flow [kgmole/h]	<empty>
Mass Flow [kg/h]	90.72
Std Ideal Liq Vol Flow [m3/h]	<empty>
Molar Enthalpy [kJ/kgmole]	<empty>
Molar Entropy [kJ/kgmole-C]	<empty>
Heat Flow [kJ/h]	<empty>
Liq Vol Flow @Std Cond [m3/h]	<empty>
Fluid Package	Basis-1
Utility Type	

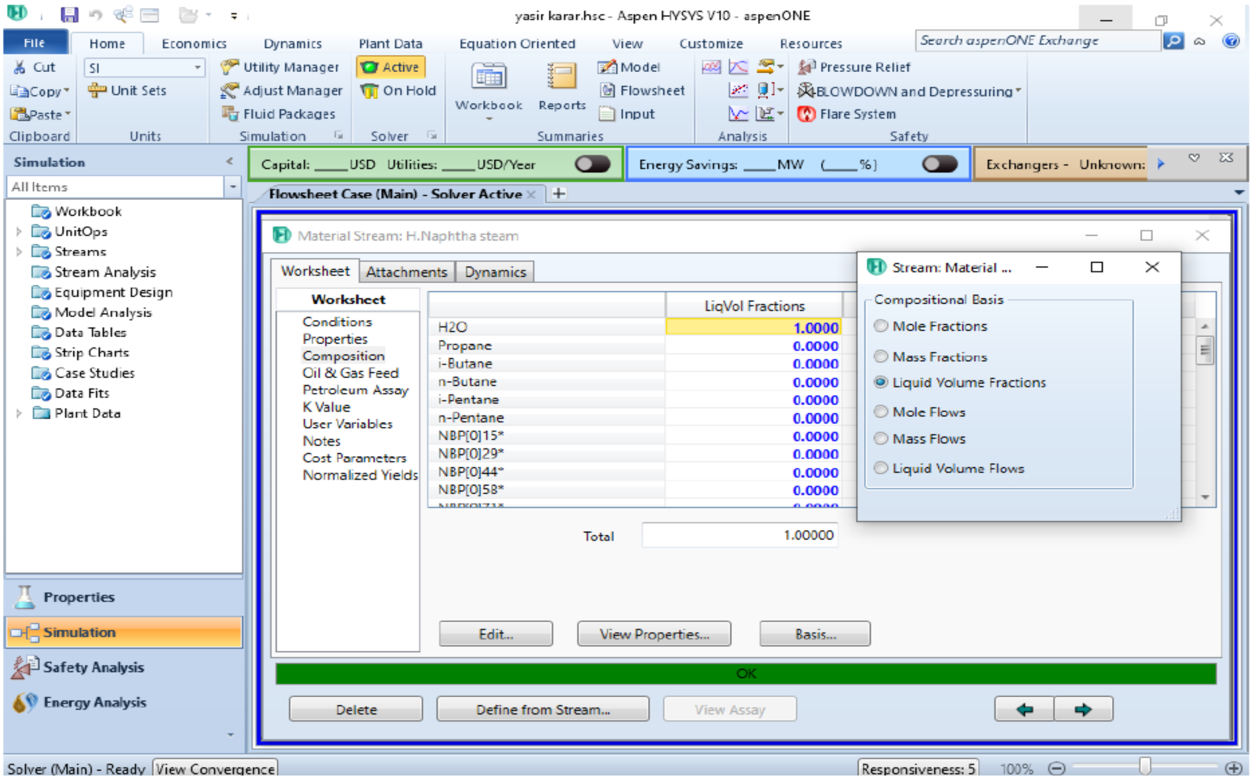
Figure 3.29: Definition of temperature, pressure and flow rate



	Mole Fractions
H2O	1
Propane	<empty>
i-Butane	<empty>
n-Butane	<empty>
i-Pentane	<empty>
n-Pentane	<empty>
NBP[0]15*	<empty>
NBP[0]29*	<empty>
NBP[0]44*	<empty>
NBP[0]58*	<empty>
NBP[0]71*	<empty>
NBP[0]84*	<empty>
NBP[0]98*	<empty>

Total 0.00000

Figure 3.30: The components are listed by mole fraction



	LiqVol Fractions
H2O	1.0000
Propane	0.0000
i-Butane	0.0000
n-Butane	0.0000
i-Pentane	0.0000
n-Pentane	0.0000
NBP[0]15*	0.0000
NBP[0]29*	0.0000
NBP[0]44*	0.0000
NBP[0]58*	0.0000
Total	1.00000

Stream: Material ...

Compositional Basis

○ Mole Fractions

○ Mass Fractions

● Liquid Volume Fractions

○ Mole Flows

○ Mass Flows

○ Liquid Volume Flows

Figure 3.31: The components are listed by liquid volume fraction

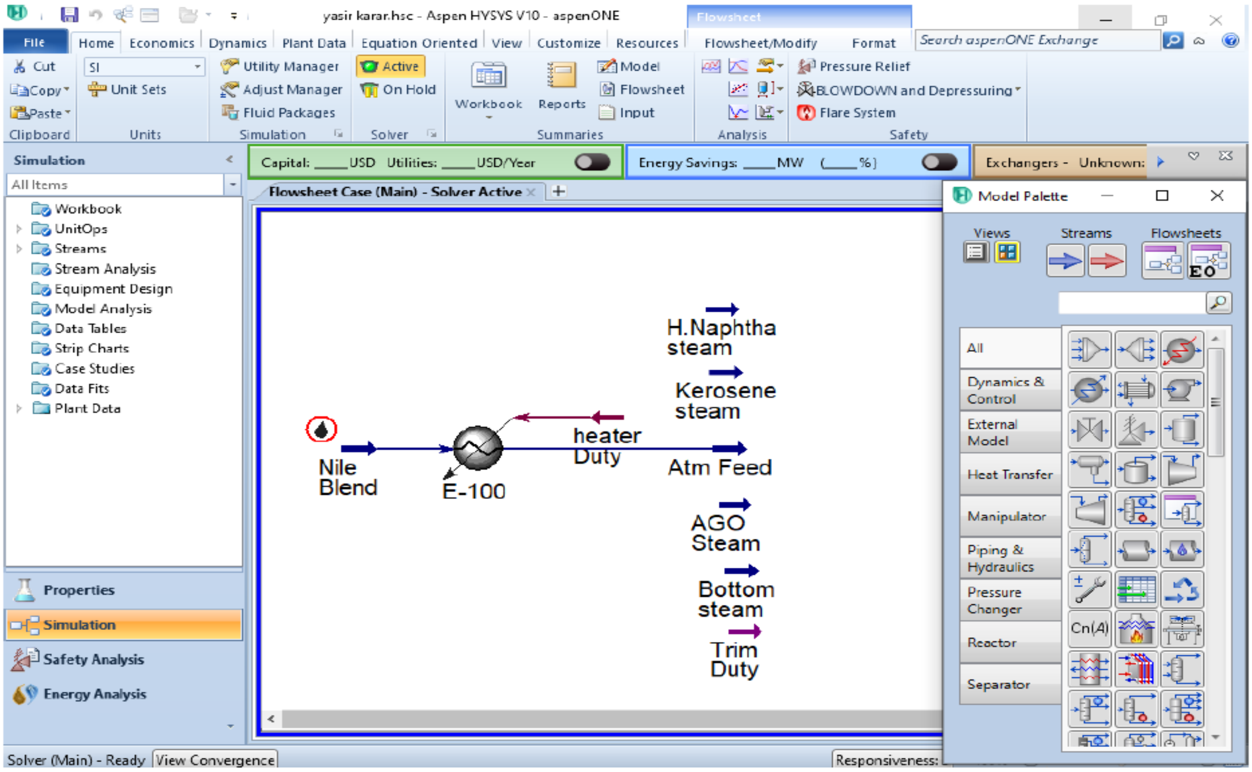


Figure 3.32: Definition of the steam streams

3.4 Atmospheric Crude Column

Atmospheric distillation towers are one of the most important pieces of equipment in the petroleum refining industry. The atmospheric tower serves to distil the crude oil into several different cuts. These include light and heavy naphtha, kerosene, Atmospheric Gas Oil and Residue.

3.4.1 Crude Distillation Unit (CDU)

The column consists of 36 stages with a partial condenser and three side strippers. The heated crude was sent to the tray 4 from bottom. Side strippers comprising 4 stages were utilized for heavy Naphtha, Kerosene and Atmospheric Gas Oil (AGO).

Fractionation was increased by reducing the partial pressures with the aid of steam (Table 3.4). There was a bottom steam entering at first tray from bottom at a rate of 707.6 kg/hr, 398.9˚C and 482.6 kPa.

Table 3.4: Side stripper specifications

Side stripper (SS)	Liquid draw stage	Vapour return stage	Stripped by	Steam flow Kg/h
Heavy Naphtha	Tray 32	Tray 34	Steam	90.72
Kerosene	Tray 22	Tray 27	Steam	90.72
AGO	Tray 13	Tray 19	Steam	131.5

description of the simulation steps for (CDU) or the procedure of the simulation by HYSYS, and also provide the procedure that’s followed to design the atmospheric distillation tower. HYSYS has a number of prebuilt column templates that is installed and customized by changing attached stream names, number of stages and default specifications, and adding side equipment. Absorber column was created on the flow sheet from model palette (Figure3.33) and the basic information were filled (Figure3.34).

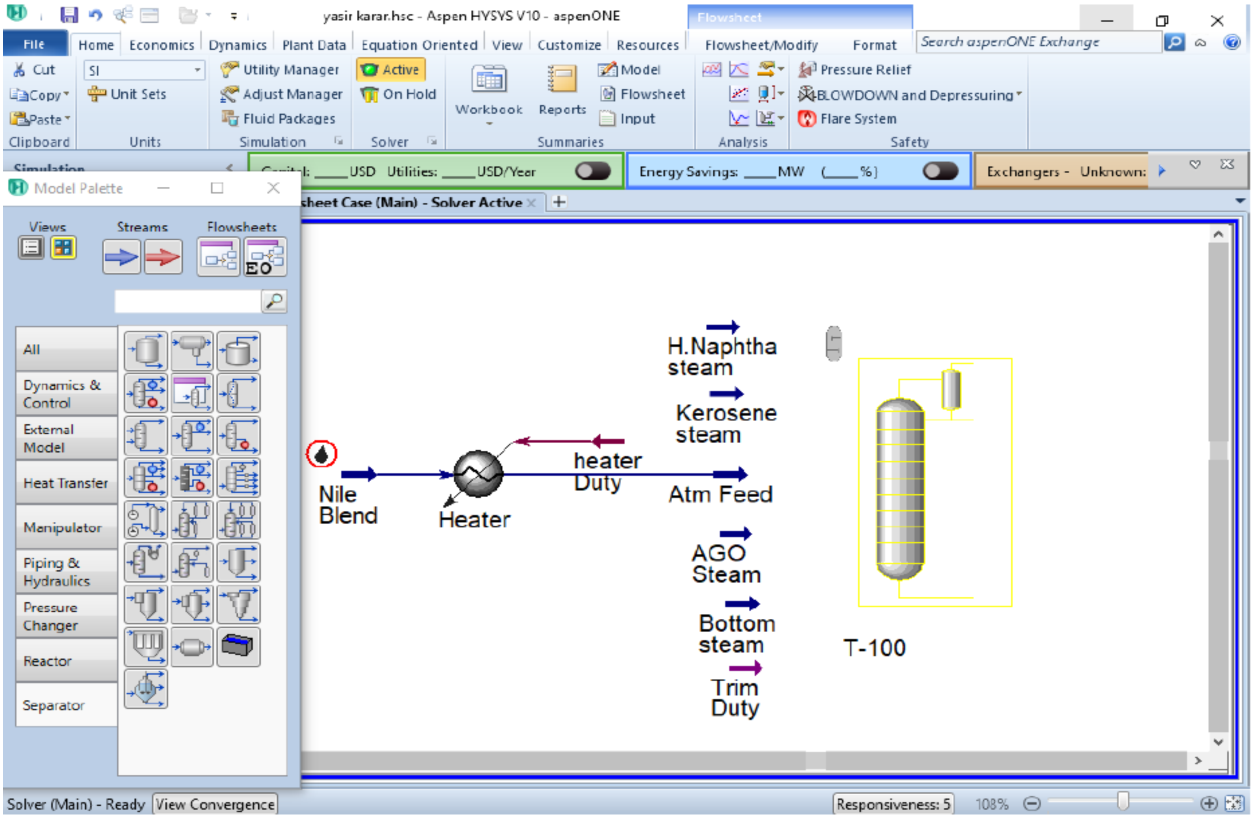
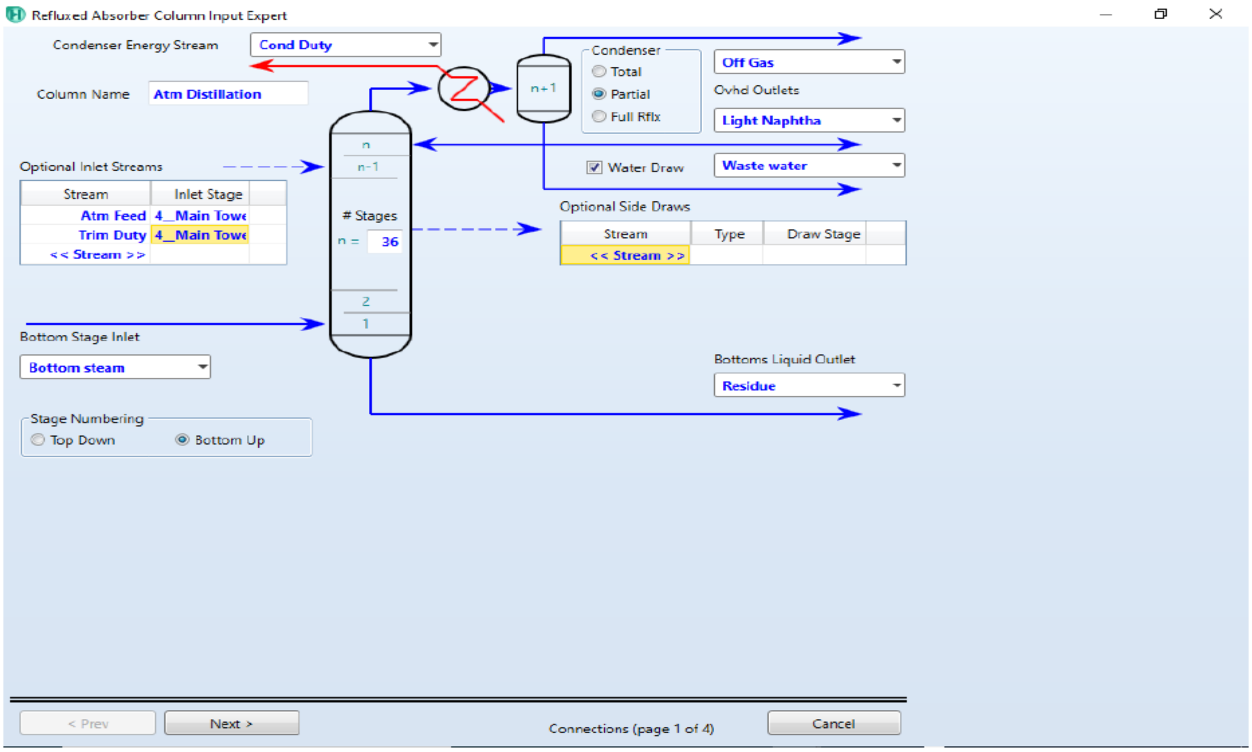


Figure 3.33: Absorber column addition



Refluxed Absorber Column Input Expert

Condenser Energy Stream: Cond Duty

Column Name: Atm Distillation

Optional Inlet Streams

Stream	Inlet Stage
Atm Feed	4__Main Towe
Trim Duty	4__Main Towe
<< Stream >>	

Condenser
○ Total
● Partial
○ Full Rflx

Ovhd Outlets: Off Gas, Light Naphtha

☑ Water Draw: Waste water

Stages n = 36

Optional Side Draws

Stream	Type	Draw Stage
<< Stream >>		

Bottom Stage Inlet: Bottom steam

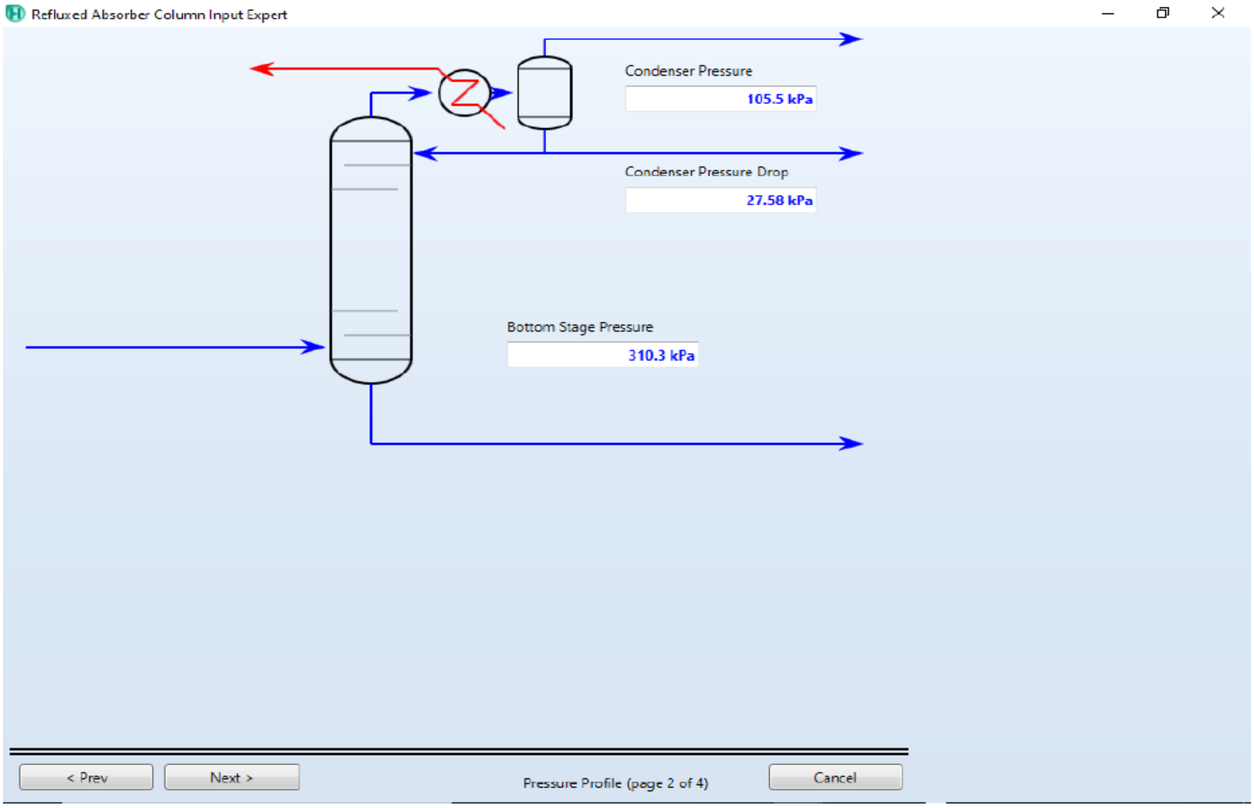
Bottoms Liquid Outlet: Residue

Stage Numbering
○ Top Down
● Bottom Up

Connections (page 1 of 4)

Figure 3.34: Absorber column input

The next button now becomes available, indicating sufficient information has been supplied to advance to the next page. Now click the Next button to advance to the pressure profile page. On the pressure profile page, specify the condenser pressure drop 27.58 kPa with a top and bottom stage pressure of 105.5kPa and 310.3kPa respectively (Figure3.35).



Condenser Pressure: 105.5 kPa

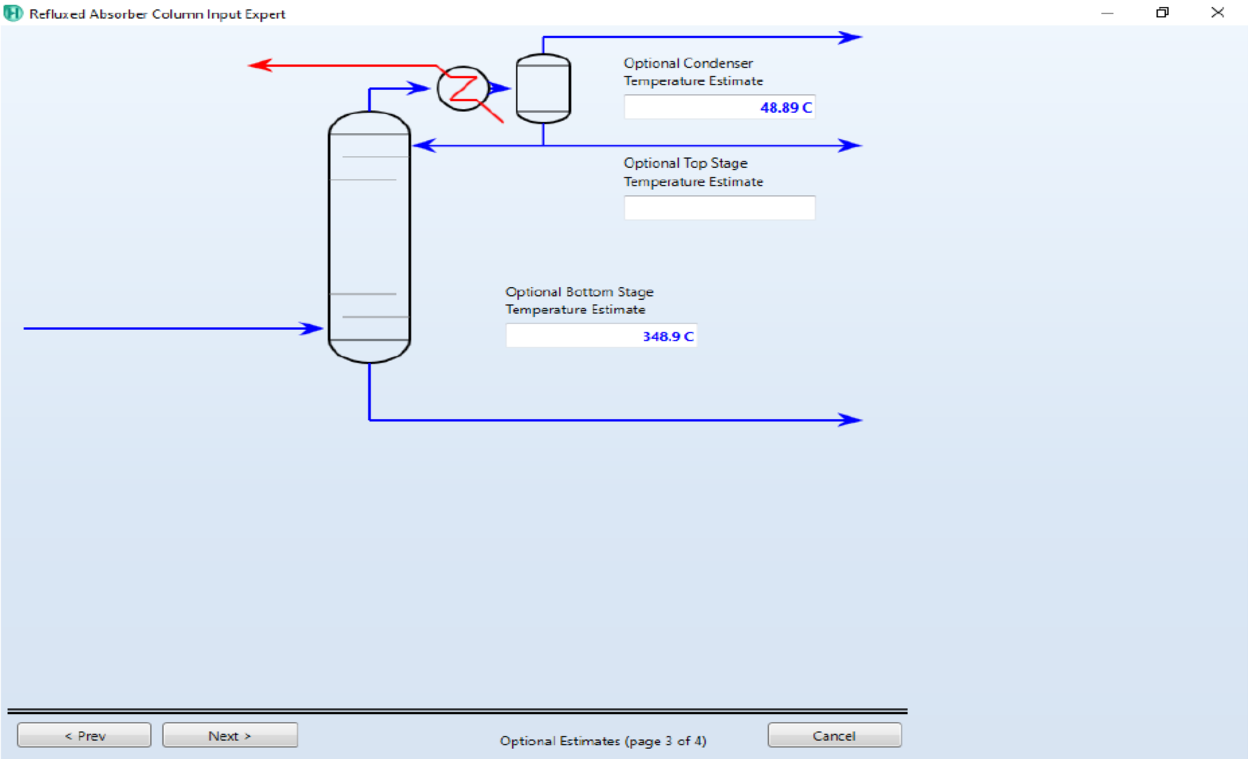
Condenser Pressure Drop: 27.58 kPa

Bottom Stage Pressure: 310.3 kPa

Figure 3.35: Pressure profile

Click the next button to advance to the optional estimates page. Although HYSYS does not usually require estimates to produce a converged column, the good estimates result in a faster solution. Specify the temperature of condenser 48.89˚C and temperature of bottom stage 348.9˚C (Figure3.36).

In general, a refluxed absorber with a partial condenser has two degrees of freedom for which HYSYS provides two default specifications. For the two specifications given, overhead vapour rate is used as an active specification, and reflux ratio as an estimate only. Specify the vapour rate equal zero, reflux ratio equal one.


Refluxed Absorber Column Input Expert

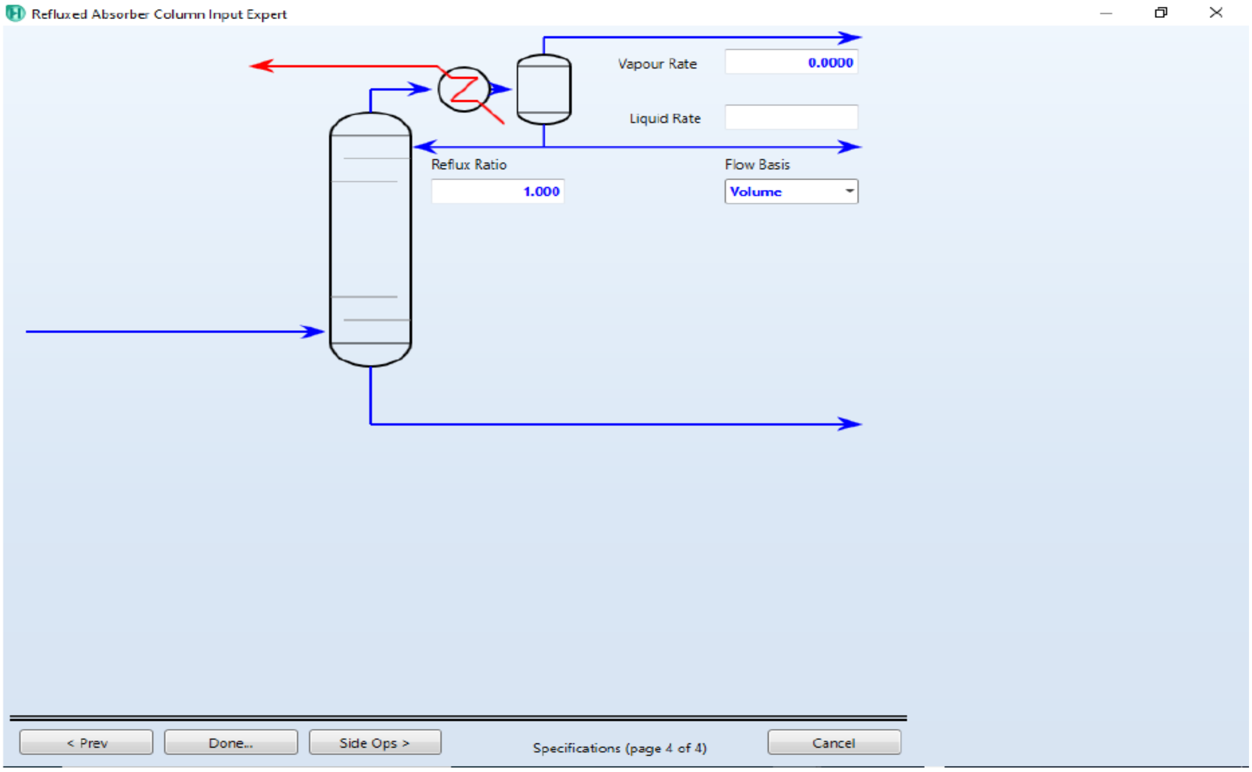
Optional Condenser Temperature Estimate: 48.89 C

Optional Top Stage Temperature Estimate:

Optional Bottom Stage Temperature Estimate: 348.9 C

Optional Estimates (page 3 of 4)

Figure 3.36: Temperature estimate


Refluxed Absorber Column Input Expert

Vapour Rate: 0.0000

Liquid Rate:

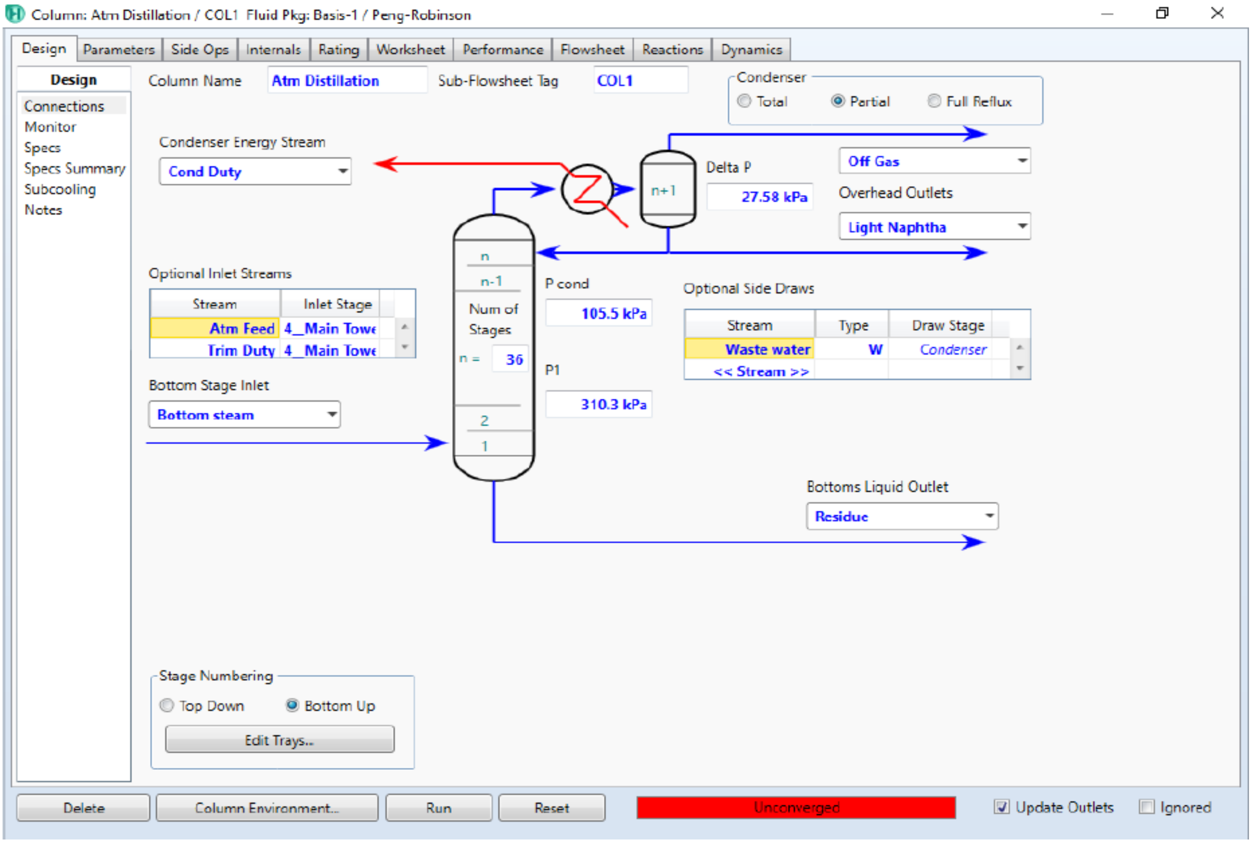
Reflux Ratio: 1.000

Flow Basis: Volume

Specifications (page 4 of 4)

Figure 3.37: Refluxed absorber with a partial condenser

Click the done button. The column property view appears on the design tab, select the monitor page. The main feature of this page is that it displays the status of column as it is being calculated, updating information with each iteration, also change specification values, and activate or deactivate specifications used by the column solver, directly from this page (Figure3.38).



Column: Atm Distillation / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Column Name: Atm Distillation
Sub-Flowsheet Tag: COL1
Condenser: ○ Total ● Partial ○ Full Reflux
Condenser Energy Stream: Cond Duty
Delta P: 27.58 kPa
Overhead Outlets: Off Gas, Light Naphtha
P cond: 105.5 kPa
P1: 310.3 kPa
Num of Stages: n = 36

Optional Inlet Streams

Stream	Inlet Stage
Atm Feed	4__Main Towe
Trim Duty	4__Main Towe

Bottom Stage Inlet: Bottom steam

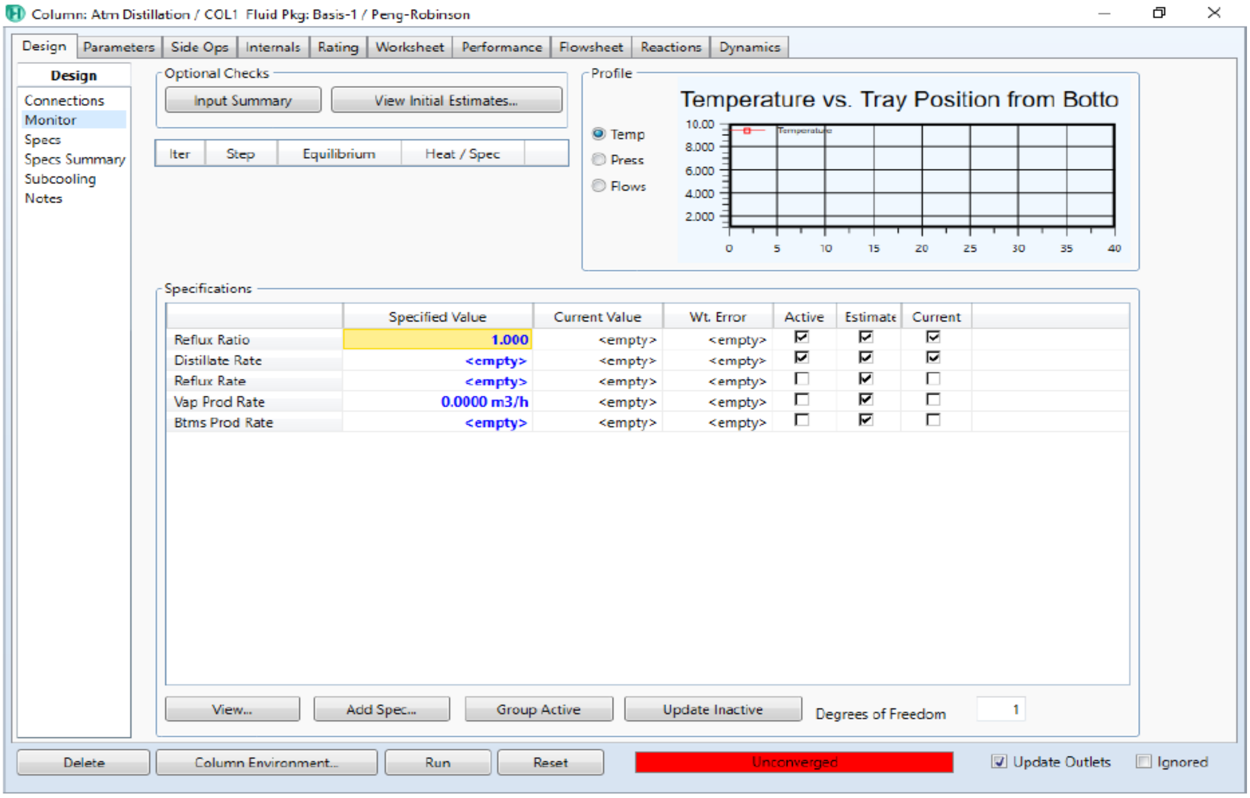
Optional Side Draws

Stream	Type	Draw Stage
Waste water	W	Condenser
<< Stream >>		

Bottoms Liquid Outlet: Residue
Stage Numbering: ○ Top Down ● Bottom Up
Unconverged
☑ Update Outlets ☐ Ignored

Figure 3.38: Installation of atmospheric distillation column

The current degrees of freedom are one (Figure3.39), indicating that only two specifications are active. Therefore, add a water draw specification called waste water rate to be the third specification active(Figure3.40). Select the specs page in the column specifications group, select column liquid flow, in the name cell, type waste water rate. No further information is required and only estimate when you run the column. In the column specifications group, click the add button. The add specs property view appears. Select column liquid flow as the column specification. Changing the name from its default to over flash, in the stage cell, condenser was select, for the over flash flowrate 3.5% from feed (Figure3.41).



	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
Reflux Ratio	1.000	<empty>	<empty>	☑	☑	☑
Distillate Rate	<empty>	<empty>	<empty>	☑	☑	☑
Reflux Rate	<empty>	<empty>	<empty>	☐	☑	☐
Vap Prod Rate	0.0000 m3/h	<empty>	<empty>	☐	☑	☐
Btms Prod Rate	<empty>	<empty>	<empty>	☐	☑	☐

Degrees of Freedom 1

Figure 3.39: Degrees of freedom

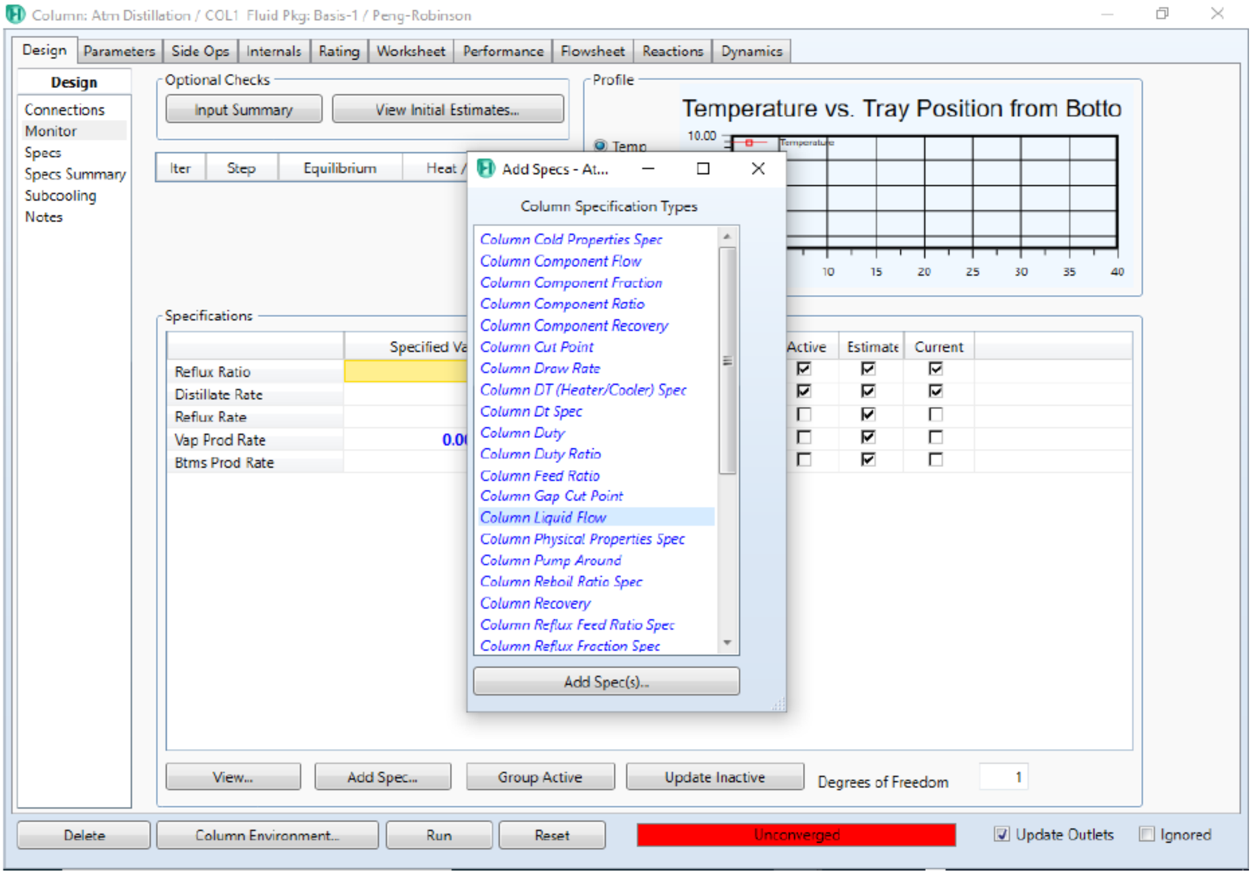
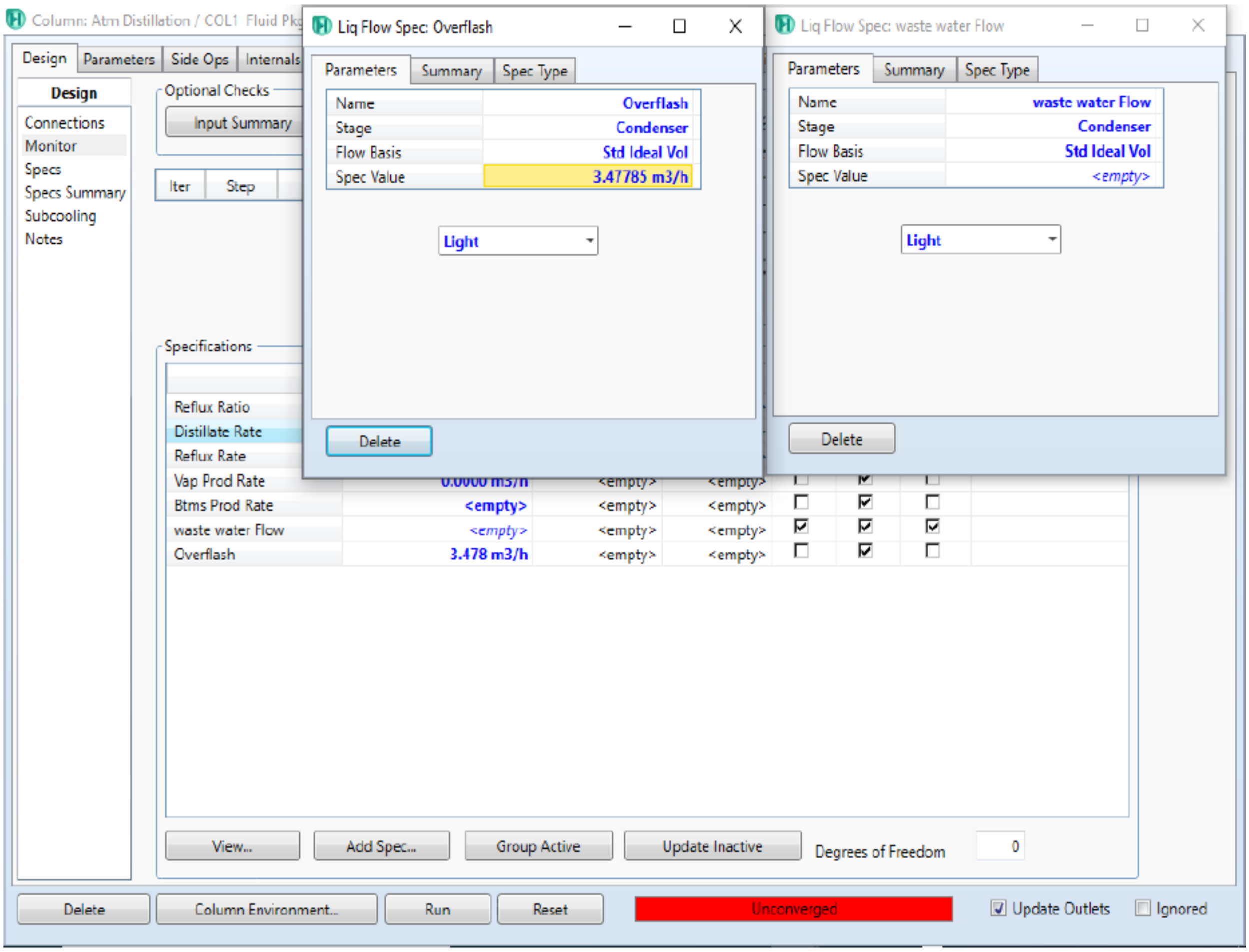


Figure 3.40: Adding specification values



Liq Flow Spec: Overflash

Name	Overflash
Stage	Condenser
Flow Basis	Std Ideal Vol
Spec Value	3.47785 m3/h

Liq Flow Spec: waste water Flow

Name	waste water Flow
Stage	Condenser
Flow Basis	Std Ideal Vol
Spec Value	<empty>

Specifications

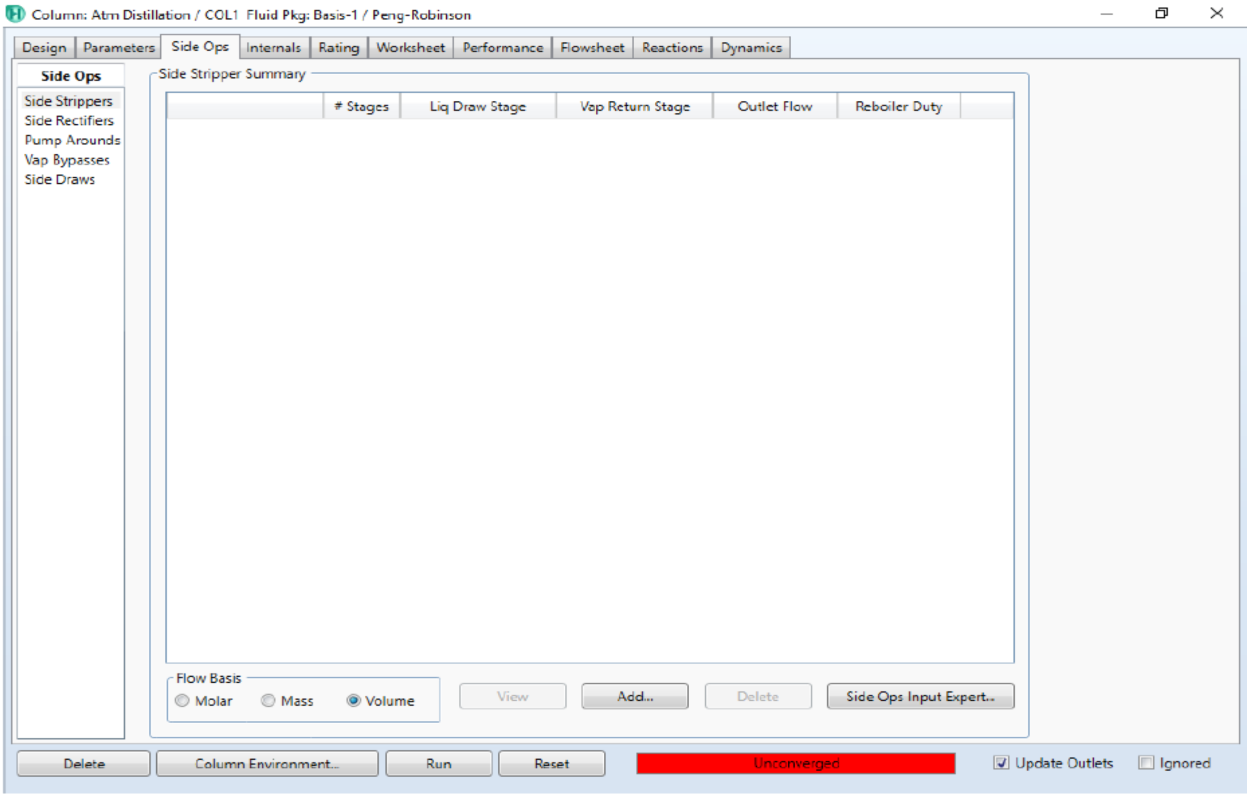
Reflux Ratio						
Distillate Rate						
Reflux Rate						
Vap Prod Rate	0.0000 m3/h	<empty>	<empty>	☐	☑	☐
Btms Prod Rate	<empty>	<empty>	<empty>	☐	☑	☐
waste water Flow	<empty>	<empty>	<empty>	☑	☑	☑
Overflash	3.478 m3/h	<empty>	<empty>	☐	☑	☐

Degrees of Freedom 0

Figure 3.41: Adding the specifications waste water and over flash

Click the side ops tab of the column property view on this tab, install, view, edit, or delete all types of side equipment. after the side strippers page were selected, click the add button (Figure 3.42). The side stripper properties view appears. In the name field, change the name to Kerosene SS and select 4-stage stripper.

In the return stage drop down list, select stage 27 (27_Main Tower). In the draw stage drop down list, select stage 22 (22_Main Tower). In the flow basis group, select the Std Ideal Vol radio button. select steam feed from configuration and add the kerosene steam, in the product stream field, enter Kerosene, the Kerosene liquid volume fraction is 0.084. For 15,000 bbl/day of crude fed to the tower, Kerosene production can be expected at 15,000*0.084 = 1,260 bbl/day (Figure 3.43).



Column: Atm Distillation / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

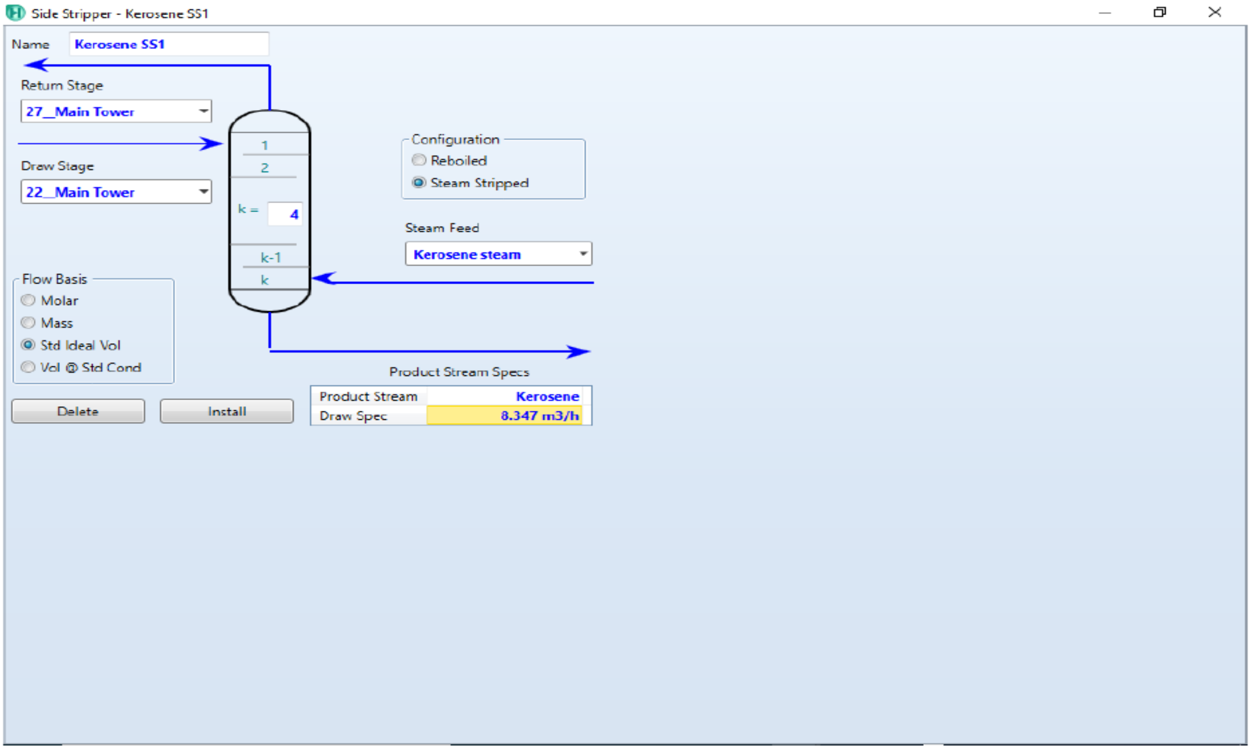
	# Stages	Liq Draw Stage	Vap Return Stage	Outlet Flow	Reboiler Duty

Flow Basis: ○ Molar ○ Mass ● Volume

Unconverged

☑ Update Outlets

Figure 3.42: Installing the side strippers



Side Stripper - Kerosene SS1

Name: Kerosene SS1

Return Stage: 27__Main Tower

Draw Stage: 22__Main Tower

k = 4

Configuration: ○ Reboiled ● Steam Stripped

Steam Feed: Kerosene steam

Flow Basis: ○ Molar ○ Mass ● Std Ideal Vol ○ Vol @ Std Cond

Product Stream Specs

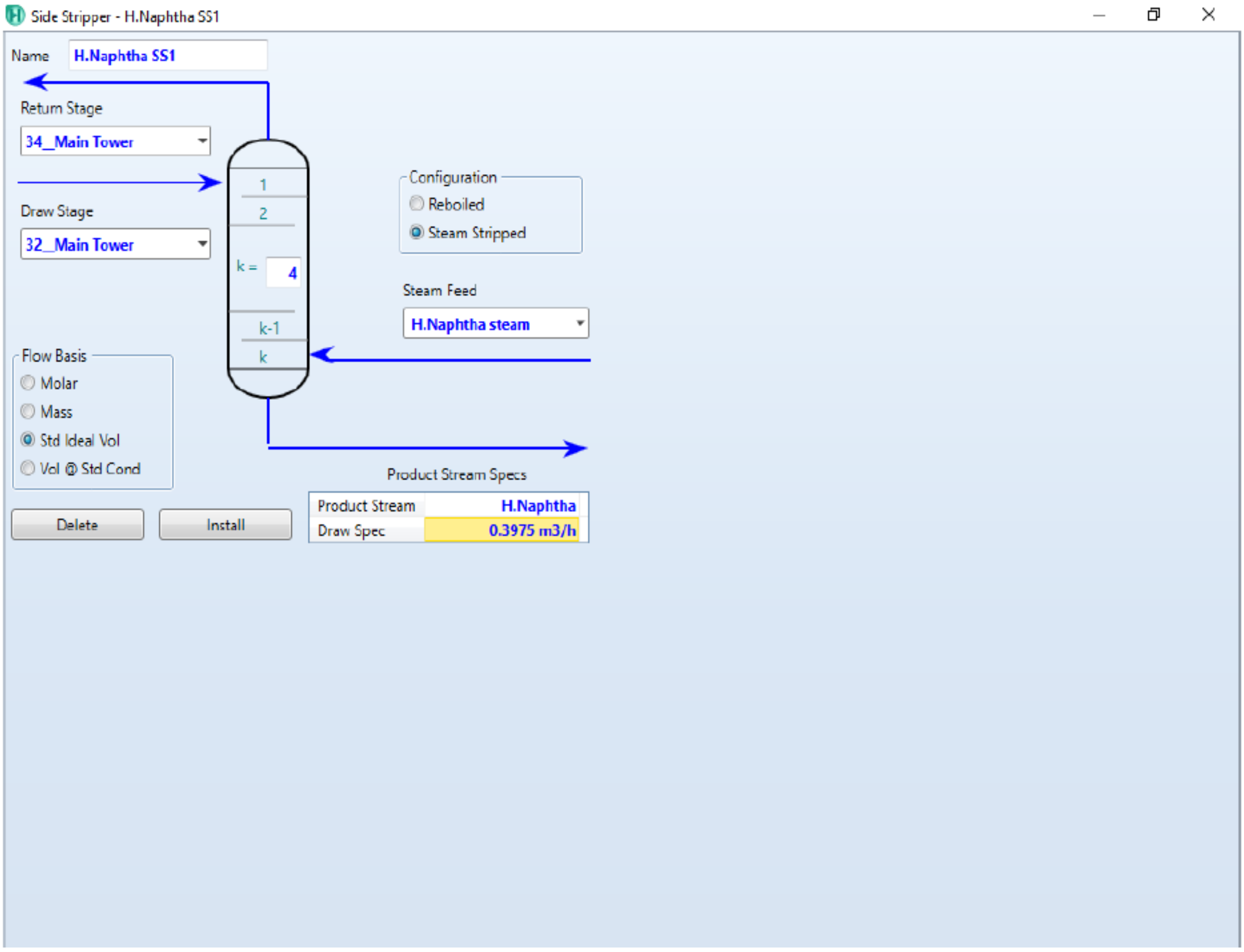
Product Stream	Kerosene
Draw Spec	8.347 m3/h

Figure 3.43: Kerosene side stripper

Use the previous steps to install the two remaining side strippers Heavy Naphtha SS and AGO SS. definition the components from the following table.

Table 3.5: Clarifies values of side strippers

Side stripper (SS)	Heavy Naphtha	AGO
Name	H.Naphtha SS	AGO SS
Liquid draw stage	Tray 32	Tray 13
Vapour return stage	Tray 34	Tray 19
Steam feed	H.Naphtha Steam	AGO Steam
Number of stage	4-stage	4-stage
Flow Basis	Std Ideal Vol	Std Ideal Vol
Product stream	H.Naphtha	AGO
Draw Specification	60 bbl/day 0.3975 m^3/h	2700bbl/day 17.89 m^3/h



Side Stripper - H.Naphtha SS1

Name: H.Naphtha SS1

Return Stage: 34__Main Tower

Draw Stage: 32__Main Tower

k = 4

Configuration
○ Reboiled
● Steam Stripped

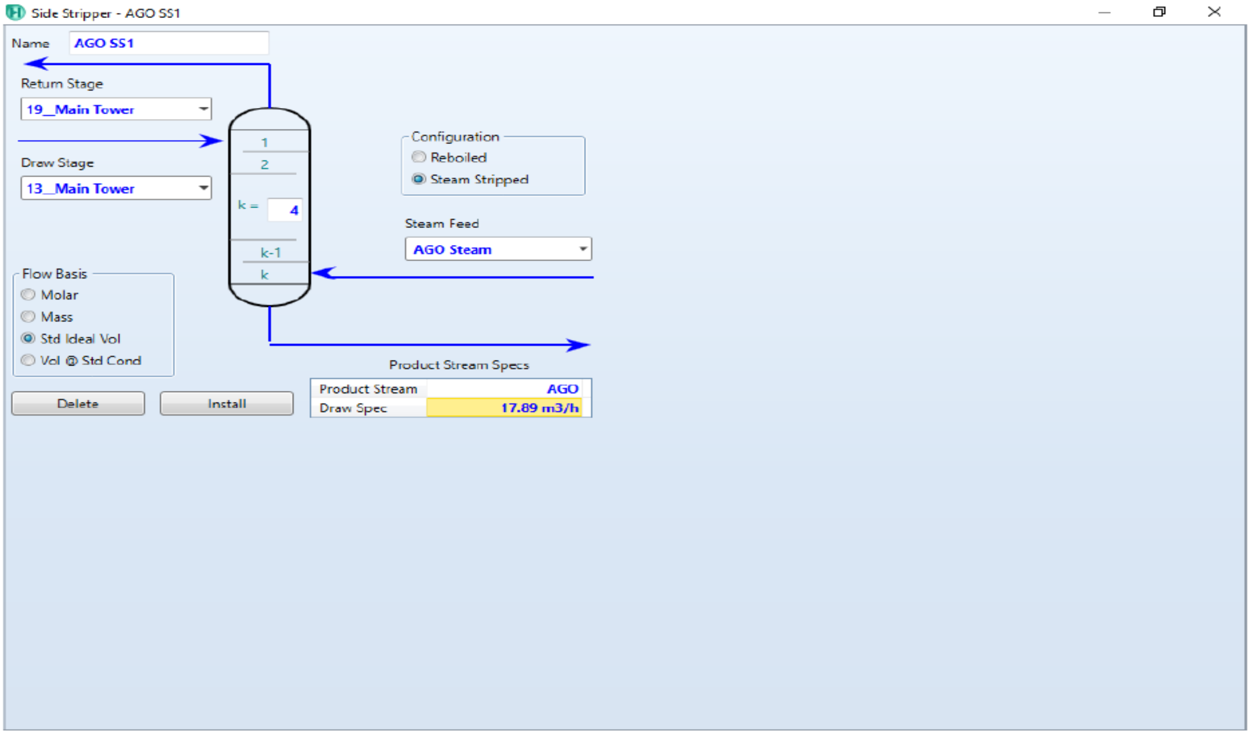
Steam Feed: H.Naphtha steam

Flow Basis
○ Molar
○ Mass
● Std Ideal Vol
○ Vol @ Std Cond

Product Stream Specs

Product Stream	H.Naphtha
Draw Spec	0.3975 m3/h

Figure 3.44: Heavy Naphtha side stripper



Name: AGO SS1

Return Stage: 19__Main Tower

Draw Stage: 13__Main Tower

k = 4

Configuration
○ Reboiled
● Steam Stripped

Steam Feed: AGO Steam

Flow Basis
○ Molar
○ Mass
● Std Ideal Vol
○ Vol @ Std Cond

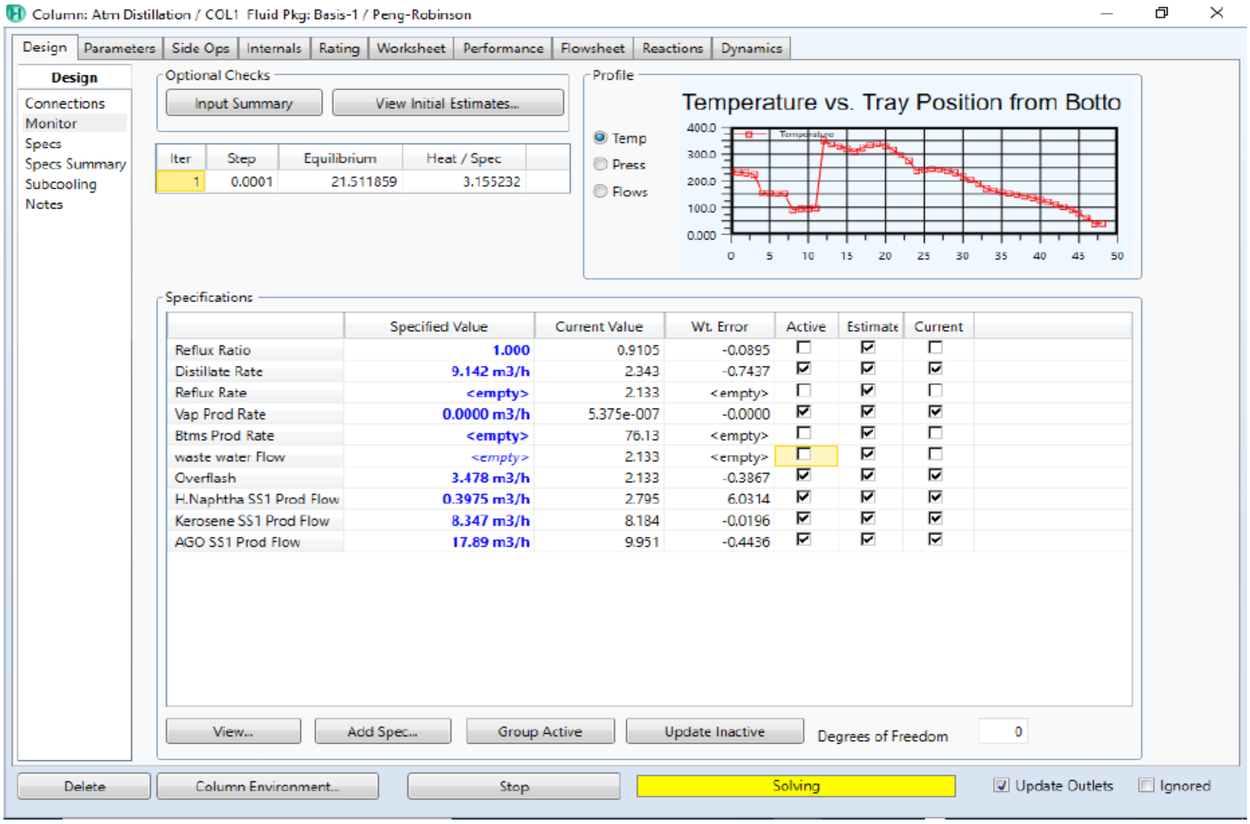
Product Stream Specs

Product Stream	AGO
Draw Spec	17.89 m3/h

Figure 3.45: Atmospheric gas oil side stripper

Select the monitor page to view the specifications matrix. The degrees of freedom is again zero, so the column is ready to be calculated, however, a flowrate for the distillate (Light Naphtha). un defined in addition, there are some specifications which are currently Active to use as estimates only, make the following final changes to the specifications, activate the overflash and vap prod rate specifications by selecting its active checkbox, deactivate the reflux ratio and waste water rate specifications(Figure 3.46). finally in the specified value cell for the distillate rate specification,the production can be expected 15,000*0.092 = 1,380 bbl/day.

HYSYS begins calculations and the information displayed on the page is updated with each iteration. The column converges as shown in the (Figure 3.47) .



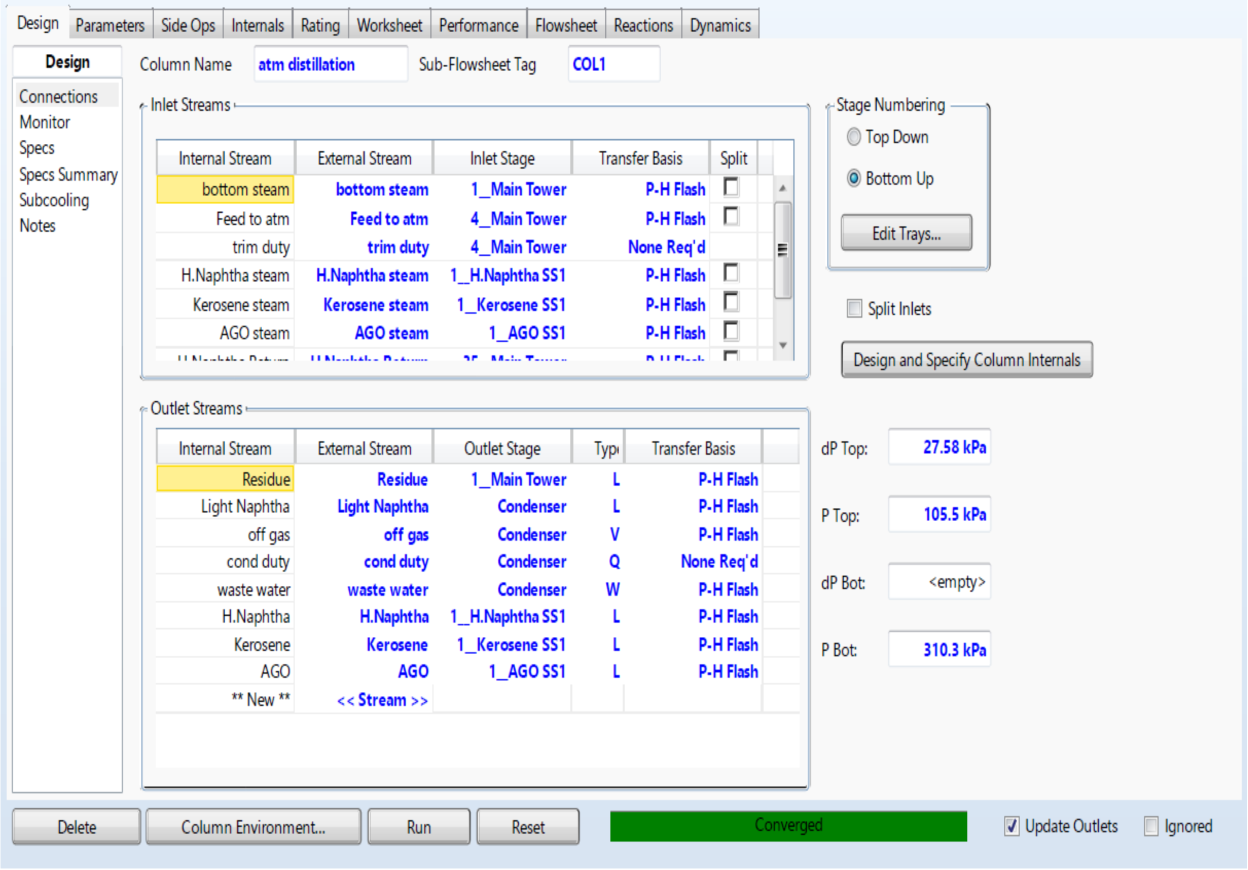
Iter	Step	Equilibrium	Heat / Spec
1	0.0001	21.511859	3.155232

Specifications

	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
Reflux Ratio	1.000	0.9105	-0.0895	☐	☑	☐
Distillate Rate	9.142 m3/h	2.343	-0.7437	☑	☑	☑
Reflux Rate	<empty>	2.133	<empty>	☐	☑	☐
Vap Prod Rate	0.0000 m3/h	5.375e-007	-0.0000	☑	☑	☑
Btms Prod Rate	<empty>	76.13	<empty>	☐	☑	☐
waste water Flow	<empty>	2.133	<empty>	☐	☑	☐
Overflash	3.478 m3/h	2.133	-0.3867	☑	☑	☑
H.Naphtha SS1 Prod Flow	0.3975 m3/h	2.795	6.0314	☑	☑	☑
Kerosene SS1 Prod Flow	8.347 m3/h	8.184	-0.0196	☑	☑	☑
AGO SS1 Prod Flow	17.89 m3/h	9.951	-0.4436	☑	☑	☑

Degrees of Freedom 0

Figure 3.46: Activate and deactivate some specifications



Column Name: atm distillation

Sub-Flowsheet Tag: COL1

Inlet Streams

Internal Stream	External Stream	Inlet Stage	Transfer Basis	Split
bottom steam	bottom steam	1__Main Tower	P-H Flash	☐
Feed to atm	Feed to atm	4__Main Tower	P-H Flash	☐
trim duty	trim duty	4__Main Tower	None Req'd	
H.Naphtha steam	H.Naphtha steam	1__H.Naphtha SS1	P-H Flash	☐
Kerosene steam	Kerosene steam	1__Kerosene SS1	P-H Flash	☐
AGO steam	AGO steam	1__AGO SS1	P-H Flash	☐

Outlet Streams

Internal Stream	External Stream	Outlet Stage	Type	Transfer Basis
Residue	Residue	1__Main Tower	L	P-H Flash
Light Naphtha	Light Naphtha	Condenser	L	P-H Flash
off gas	off gas	Condenser	V	P-H Flash
cond duty	cond duty	Condenser	Q	None Req'd
waste water	waste water	Condenser	W	P-H Flash
H.Naphtha	H.Naphtha	1__H.Naphtha SS1	L	P-H Flash
Kerosene	Kerosene	1__Kerosene SS1	L	P-H Flash
AGO	AGO	1__AGO SS1	L	P-H Flash
** New **	<< Stream >>			

dP Top: 27.58 kPa

P Top: 105.5 kPa

dP Bot: <empty>

P Bot: 310.3 kPa

Figure 3.47: Running the distillation tower

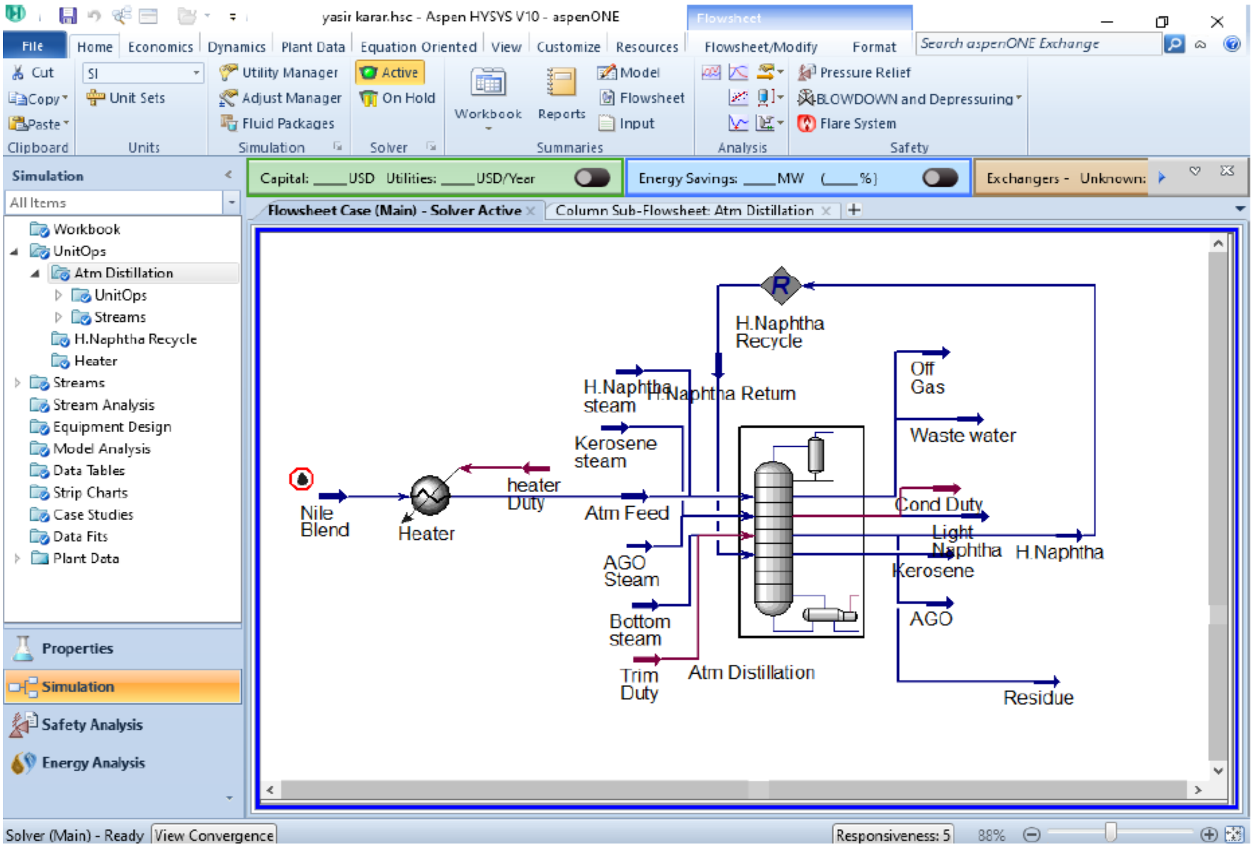


Figure 3.48: Atmospheric distillation tower

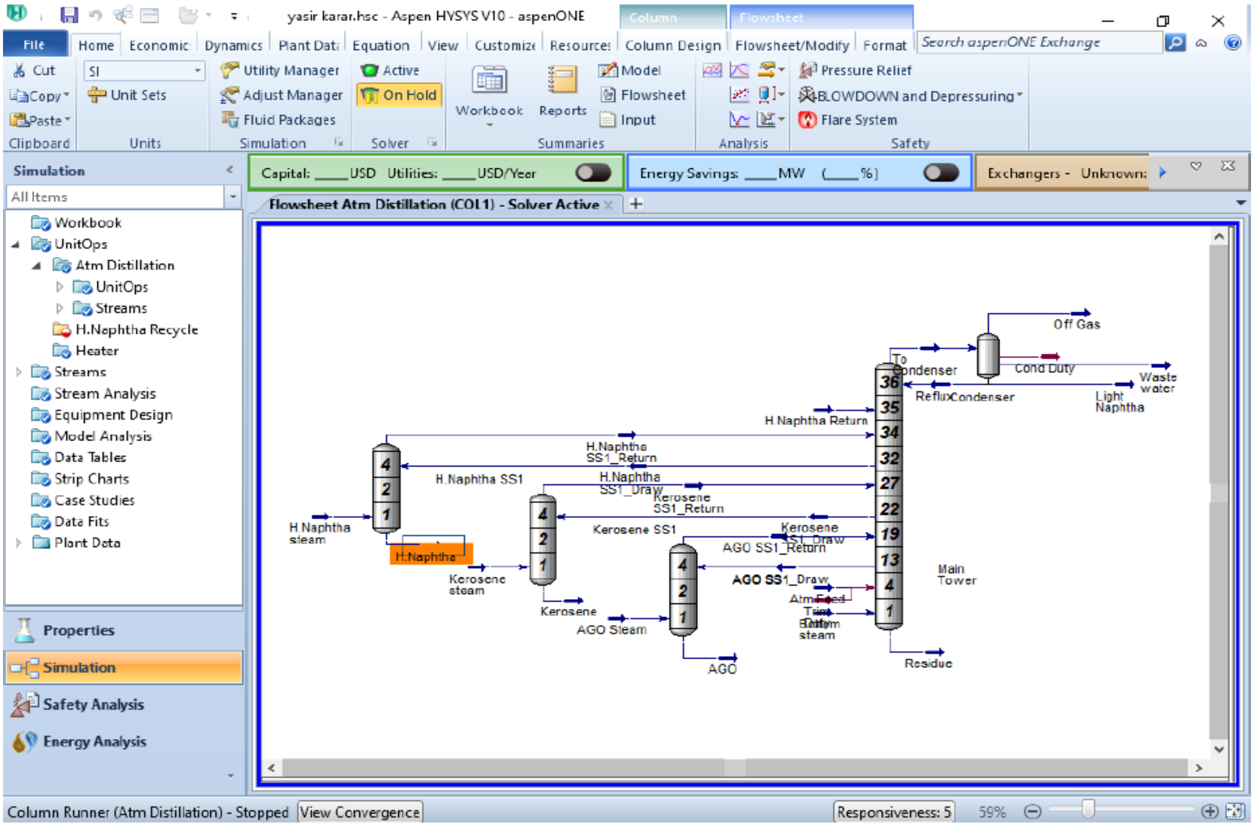


Figure 3.49: Atmospheric distillation column

In El Obeid refinery the amount of inputs is small, because the capacity of the atmospheric distillation tower is 15,000 barrels per day, and to increase the productivity of light products, to increase the amount of crude oil entering the atmospheric distillation tower and design a vacuum distillation tower, therefore, it was proposed to add a second atmospheric distillation tower with identical specifications to the first tower to increase the residue that is as feed into the vacuum distillation tower and the design shape becomes as follows (Figure3.50).

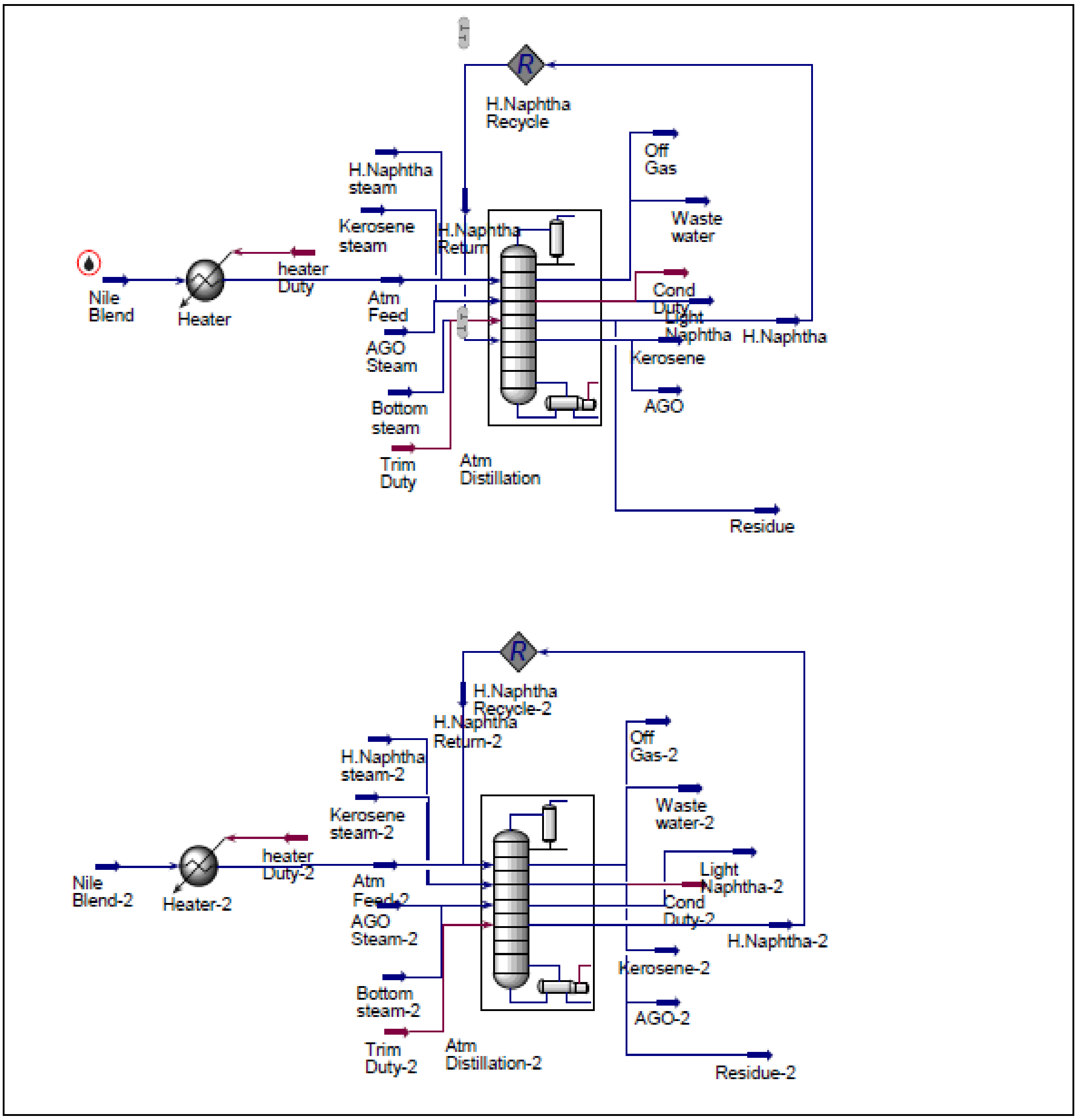


Figure3.50: Two atmospheric distillation towers

3.4.2 Vacuum Distillation Unit (VDU)

Vacuum distillation is a part of the refining process that helps to produce petroleum products out of the heavier oils left over from atmospheric distillation, vacuum distillation furnace may be classified into two types, wet and dry. In wet type, steam is injected into the furnace coils and that helps to lower the partial pressure of feed as well as steam carries the feed vapours through the furnace tube more rapidly. In dry type, steam injection is not done in the furnace. Steam injection lowers the steam consumption in the vacuum ejector systems. The first step of the designed VDU is to mix the bottom residue from the atmospheric distillation column with the vapour stream to increase velocity and minimize coke formation within the heater, afterward it is heated in a furnace to match the vacuum column conditions and controlling temperature of the furnace by MATLAB, then feed to vacuum column. Which makes start by choosing the mixer from the model palette list, as shown in (Figure .51).

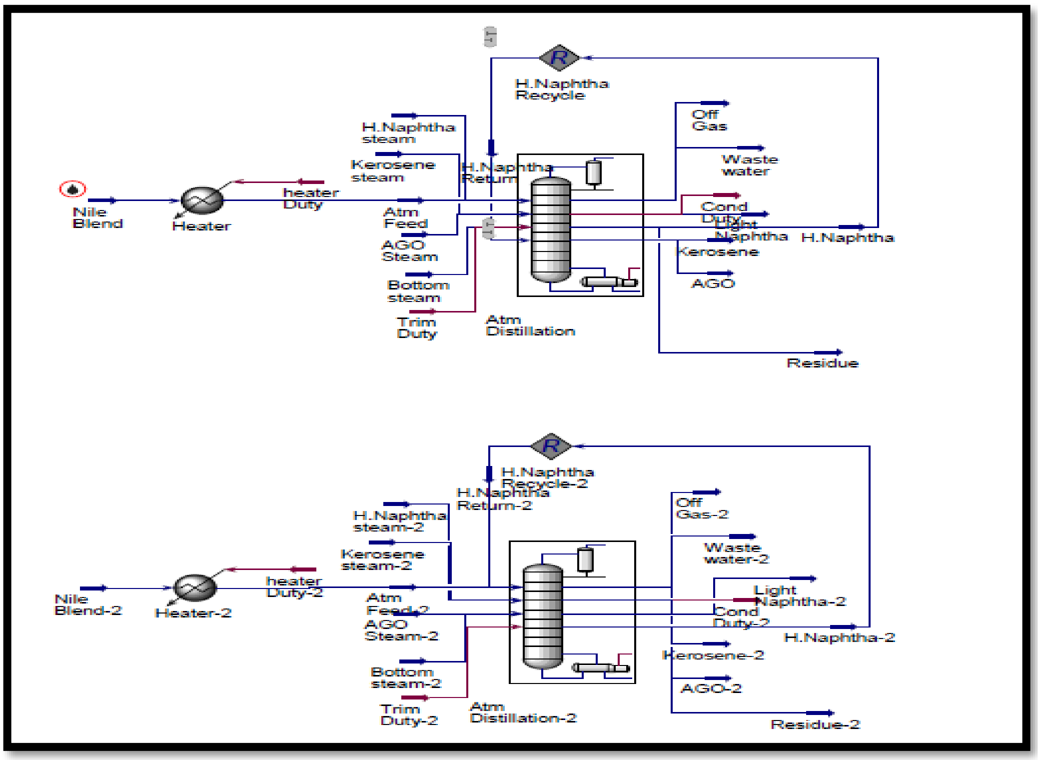


Figure3.51: Mixer selection from the model palette

After the mixer is selected on the scheme. Determine the inlet steam stream to mix it with the remaining two atmospheric distillation towers as shown in (Figure 3.52), (Figure 3.54), (Figure 3.55). Double click the material stream icon on the model palette. In the stream property view, click in the stream name cell and rename the stream steam and define the conditions (Temperature, Pressure and Flowrate) of this stream (Figure 3.53).

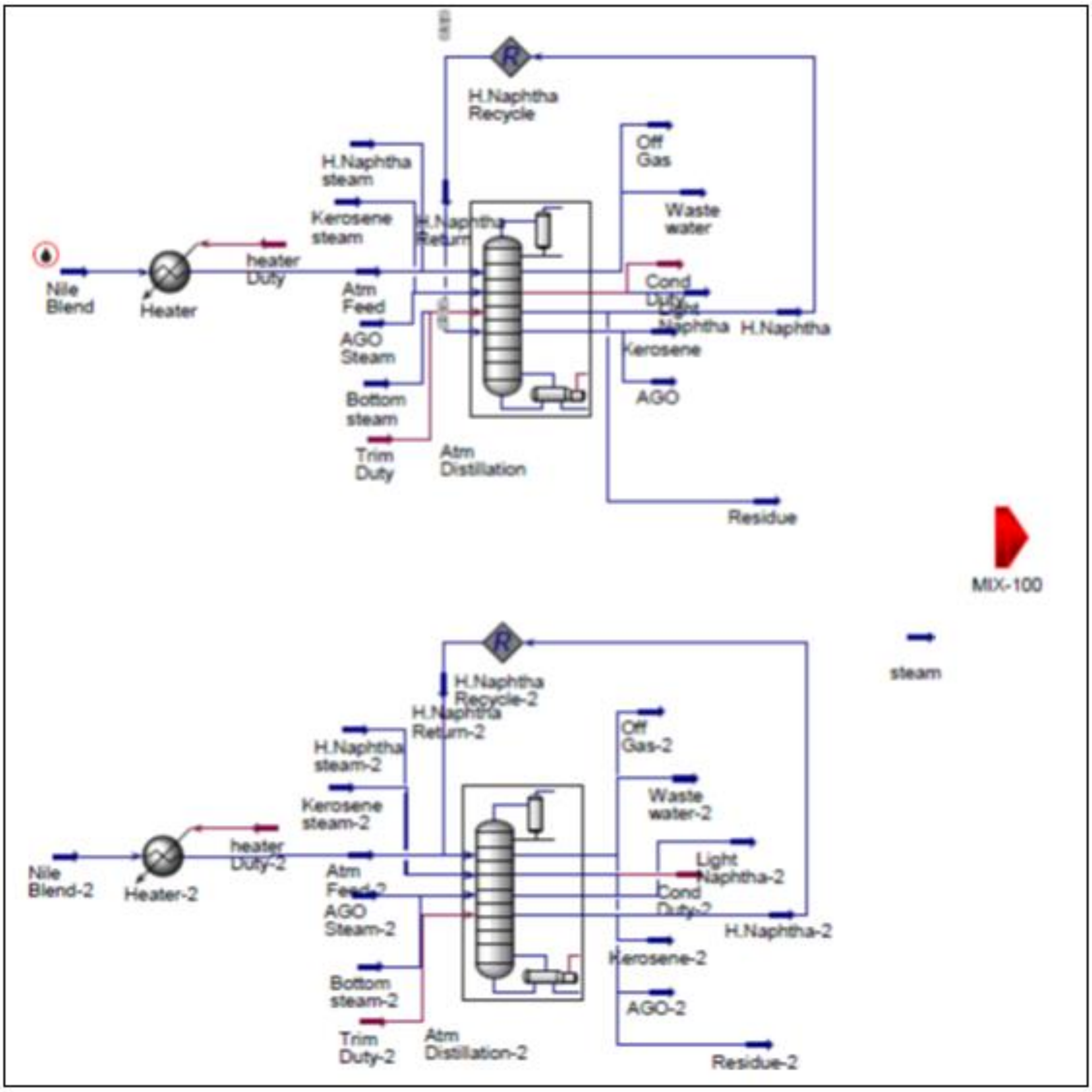
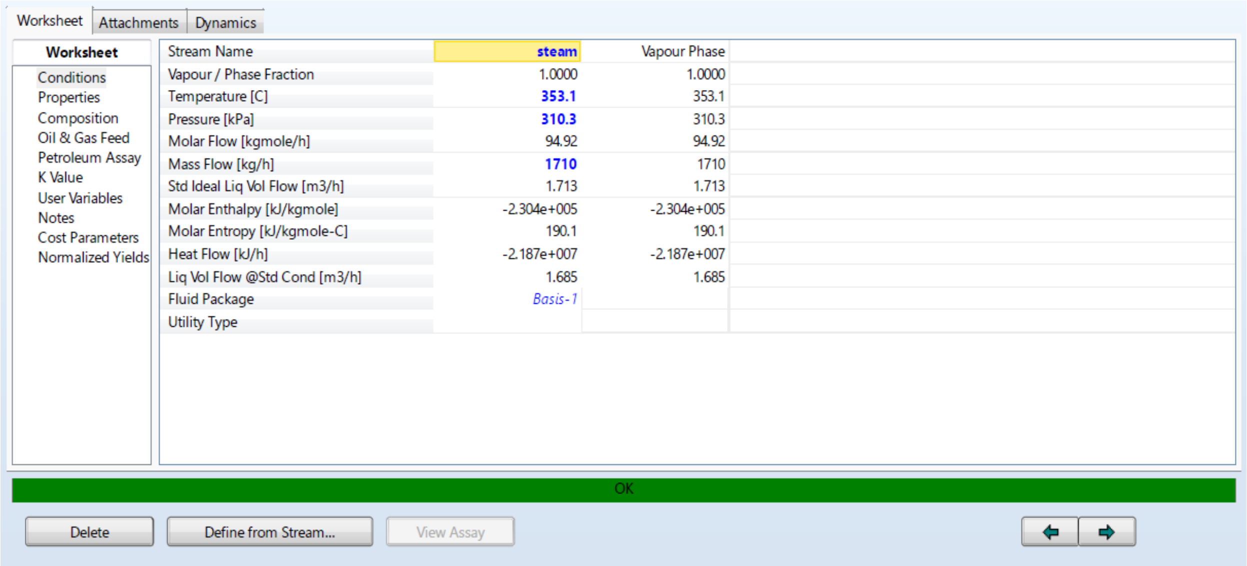


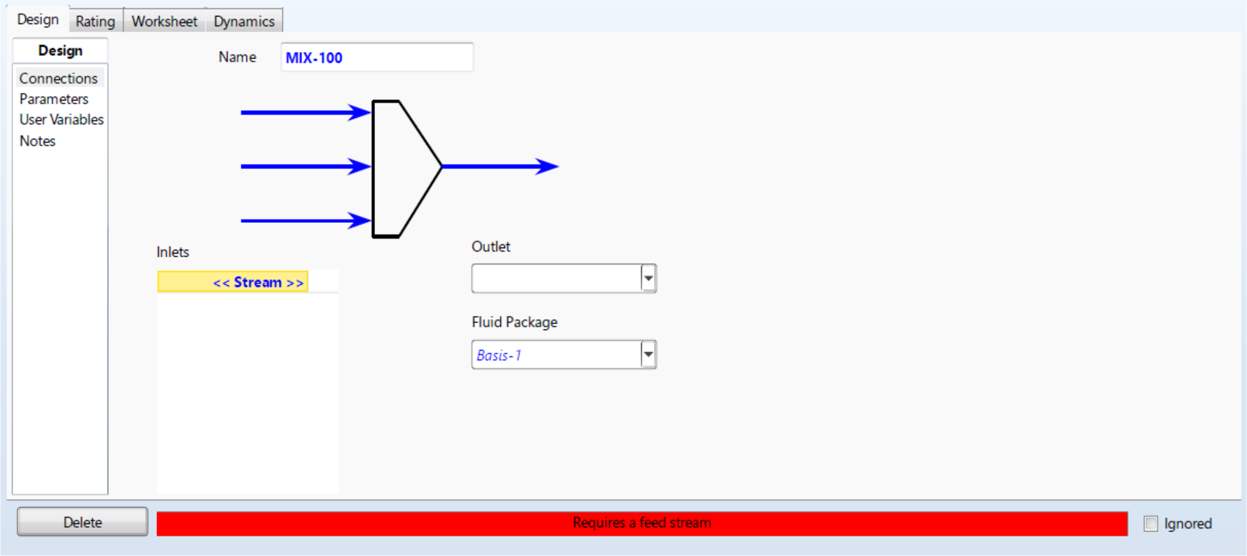
Figure3.52: The mixer on flowsheet

Press on icon MIX-100 to change name and addition Three inlet streams (Residue, Residue-2, Steam), one out let stream (crude oil) as show as (Figure3.55).



Stream Name	steam	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	353.1	353.1
Pressure [kPa]	310.3	310.3
Molar Flow [kgmole/h]	94.92	94.92
Mass Flow [kg/h]	1710	1710
Std Ideal Liq Vol Flow [m3/h]	1.713	1.713
Molar Enthalpy [kJ/kgmole]	-2.304e+005	-2.304e+005
Molar Entropy [kJ/kgmole-C]	190.1	190.1
Heat Flow [kJ/h]	-2.187e+007	-2.187e+007
Liq Vol Flow @Std Cond [m3/h]	1.685	1.685
Fluid Package	Basis-1	
Utility Type		

Figure3.53: Definition of steam conditions



Name MIX-100

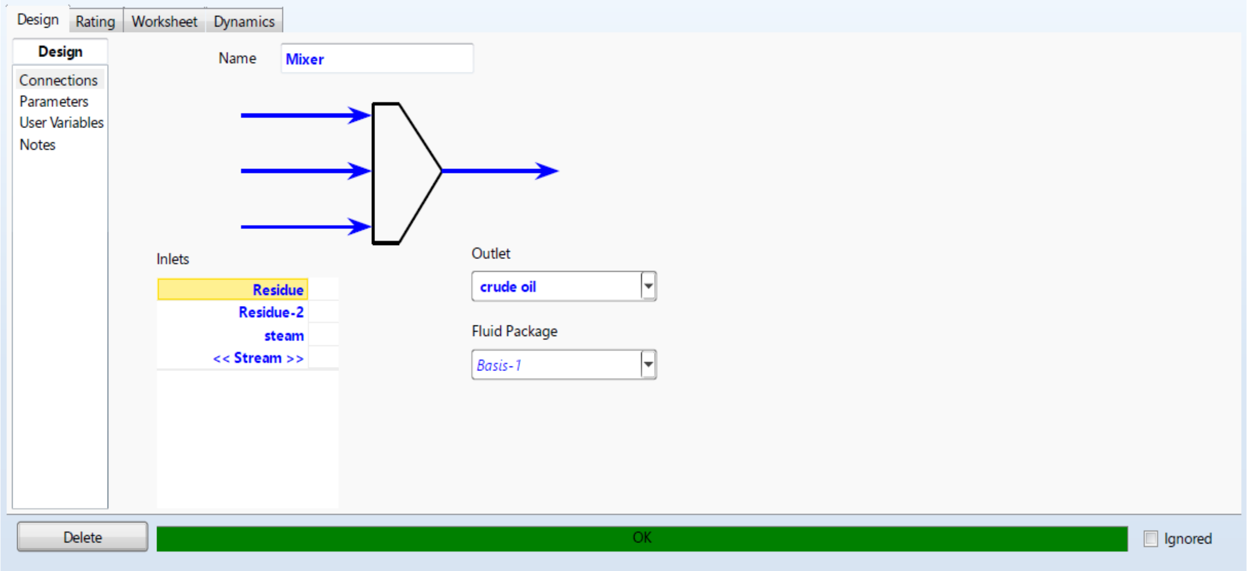
Inlets
<< Stream >>

Outlet

Fluid Package
Basis-1

Requires a feed stream

Figure3.54: Mixer without feed stream



Name Mixer

Inlets
Residue
Residue-2
steam
<< Stream >>

Outlet
crude oil

Fluid Package
Basis-1

OK

Figure3.55: Mixer inlet and outlet

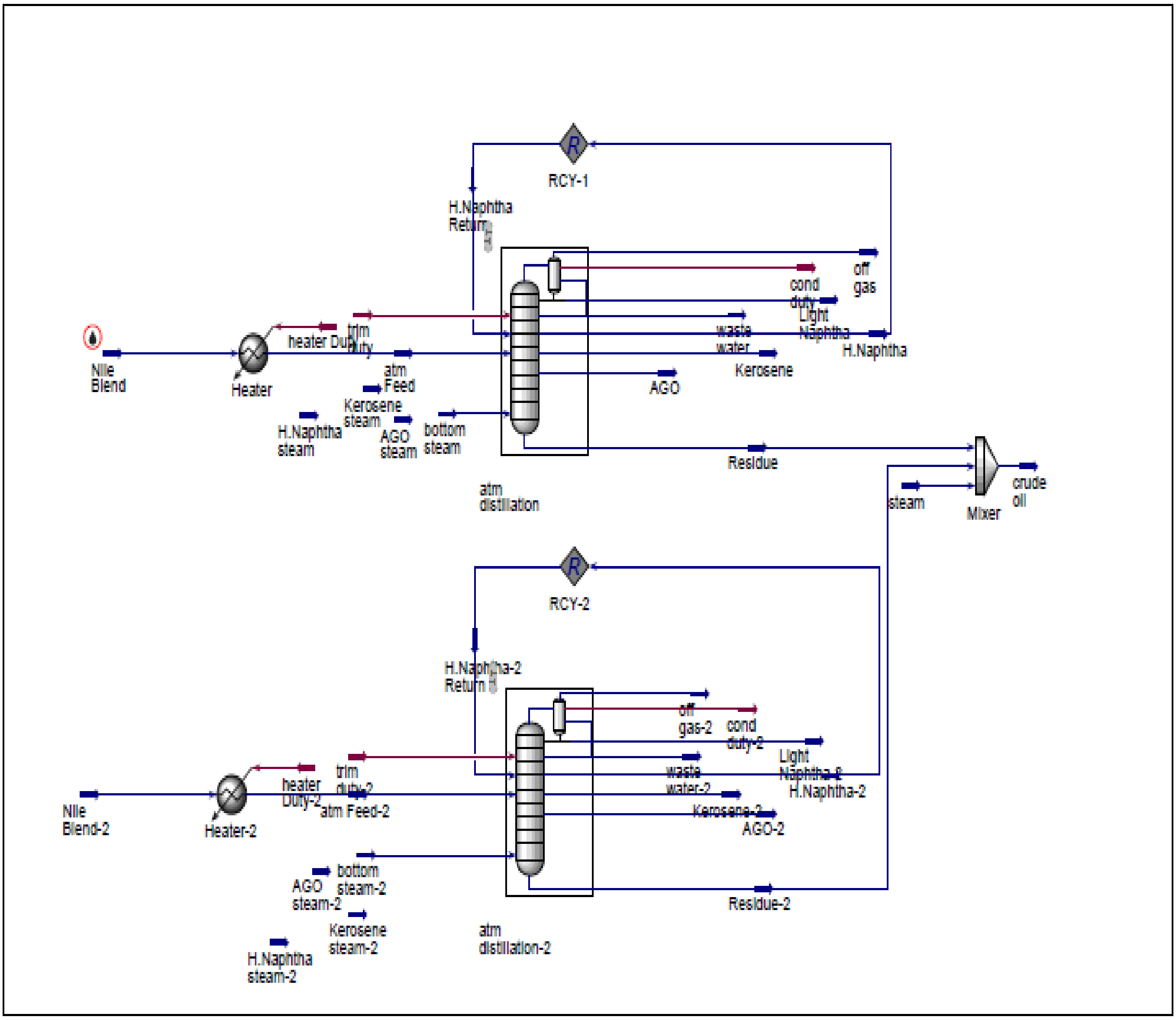


Figure3.56: Mixer after definition

Secondly, by creating a new heater from the model palette list, as shown in (Figure 3.57). press on icon E-100 (Figure3.58) to change name (vacuum heater) and addition inlet stream (crude oil), out let (Feed to vacuum), energy stream (Heater duty 3) as show as (Figure3.59). after that we can define two specifications from worksheet (temperature, pressure) for stream (Feed to vacuum) and control the temperature of the crude using the flow of the furnace fuel using MATLAB program V.7.1 as shown in (Figure3.60).

Hysys program calculated all the specifications for this stream(Figure3.61)

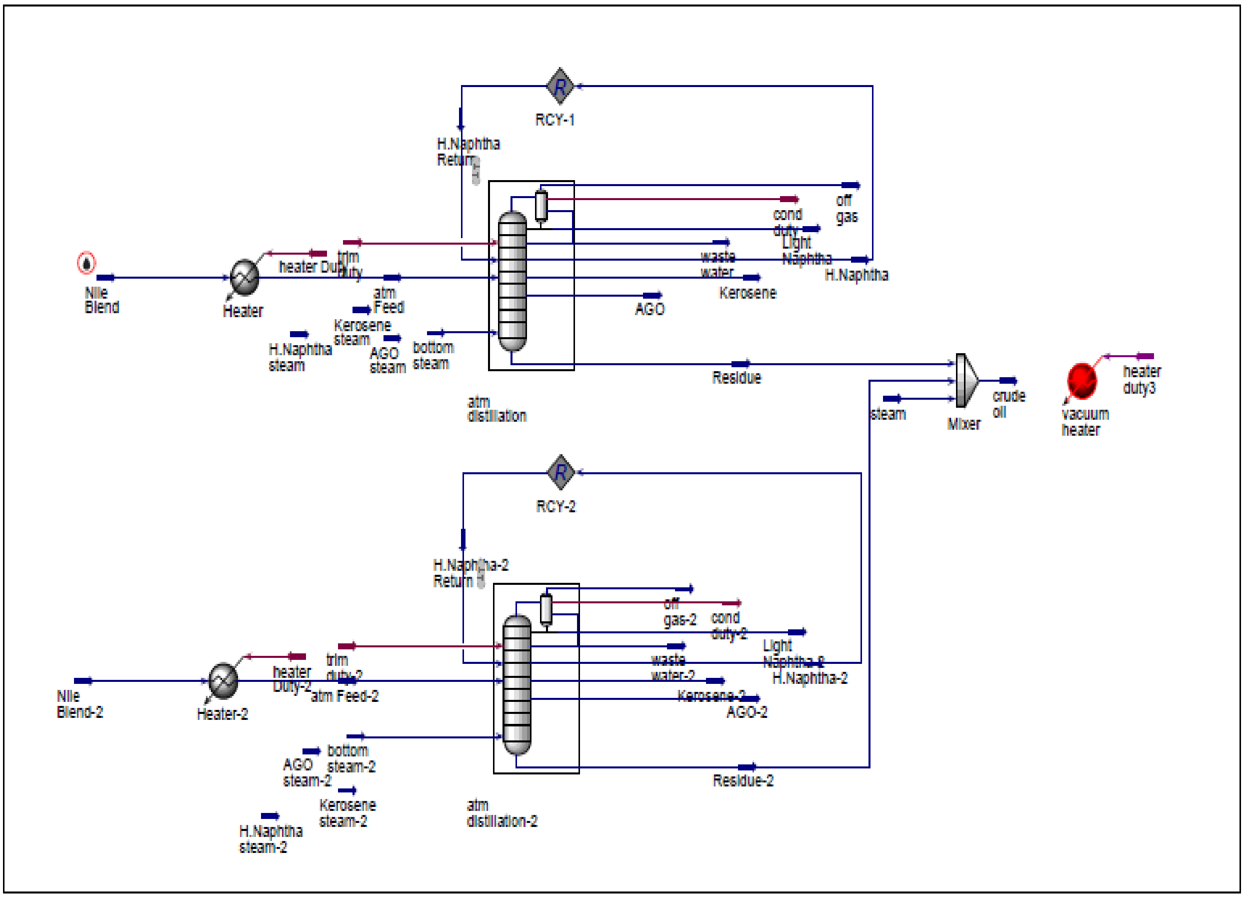
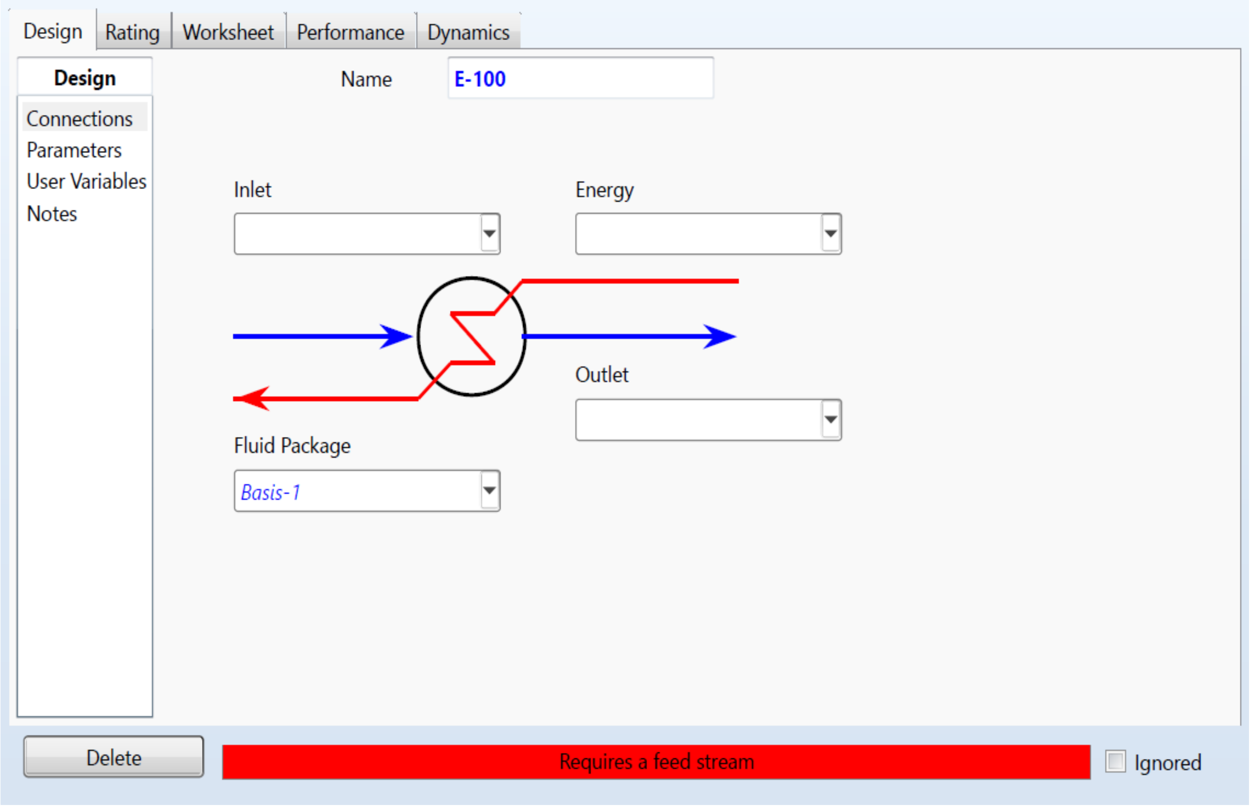


Figure3.57: Heater selection

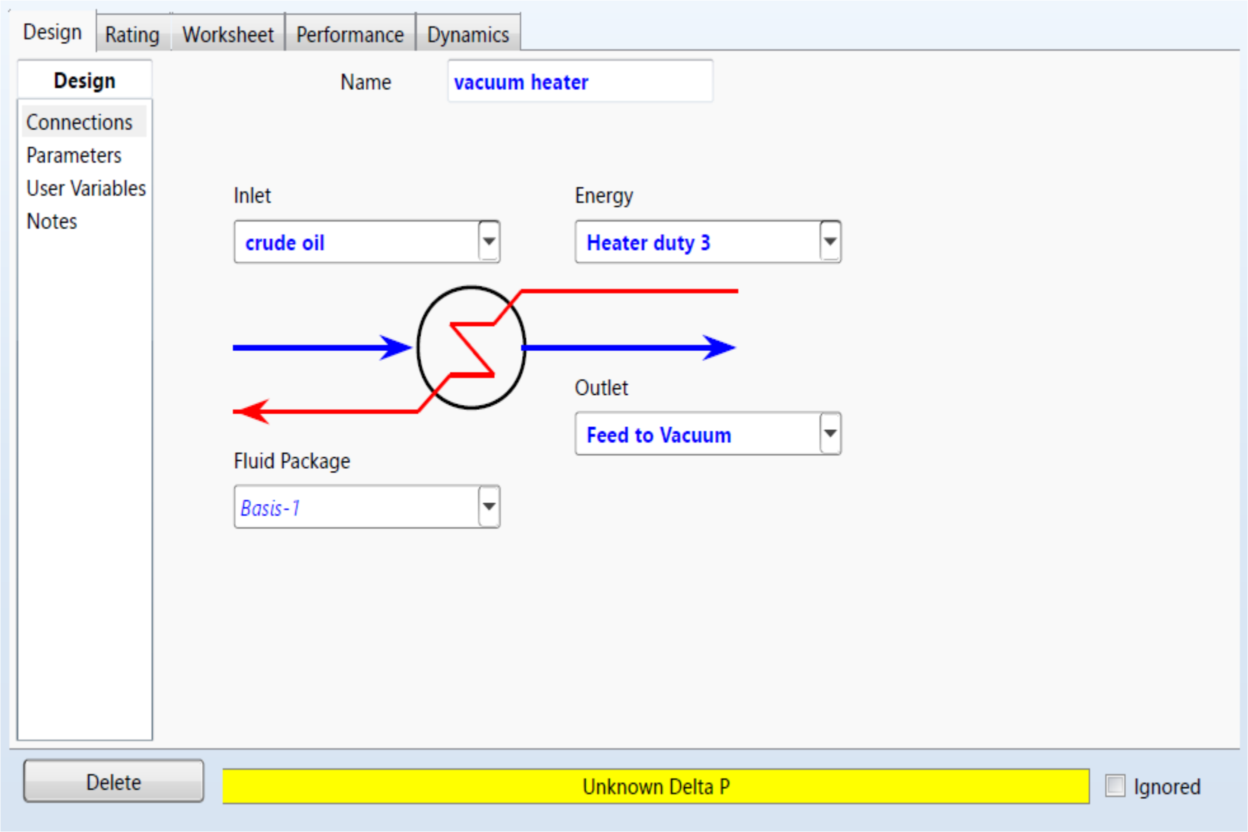


Name E-100

Fluid Package Basis-1

Requires a feed stream

Figure3.58: The Heater before feed stream



Name: vacuum heater

Inlet: crude oil

Energy: Heater duty 3

Outlet: Feed to Vacuum

Fluid Package: Basis-1

Unknown Delta P

☐ Ignored

Figure3.59: The Heater after definition

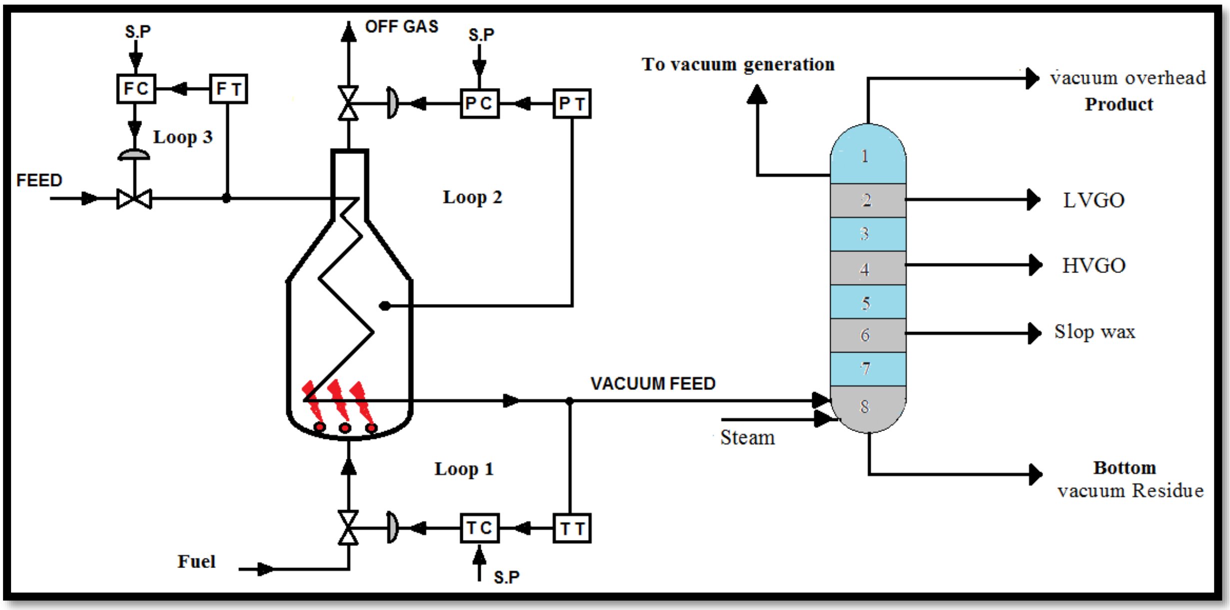
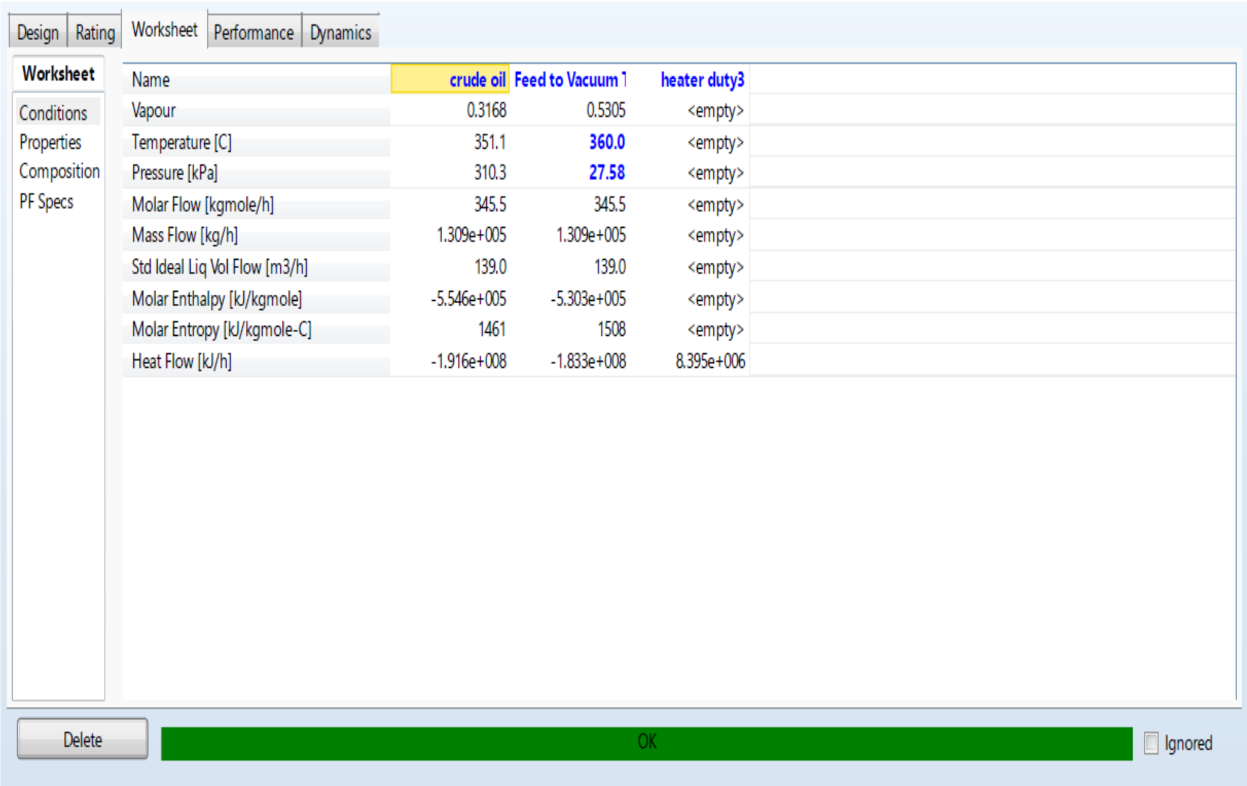


Figure 3.60: Control strategy of the furnace



Name	crude oil	Feed to Vacuum 1	heater duty3
Vapour	0.3168	0.5305	<empty>
Temperature [C]	351.1	360.0	<empty>
Pressure [kPa]	310.3	27.58	<empty>
Molar Flow [kgmole/h]	345.5	345.5	<empty>
Mass Flow [kg/h]	1.309e+005	1.309e+005	<empty>
Std Ideal Liq Vol Flow [m3/h]	139.0	139.0	<empty>
Molar Enthalpy [kJ/kgmole]	-5.546e+005	-5.303e+005	<empty>
Molar Entropy [kJ/kgmole-C]	1461	1508	<empty>
Heat Flow [kJ/h]	-1.916e+008	-1.833e+008	8.395e+006

Figure3.61: Temperature and pressure definition for stream

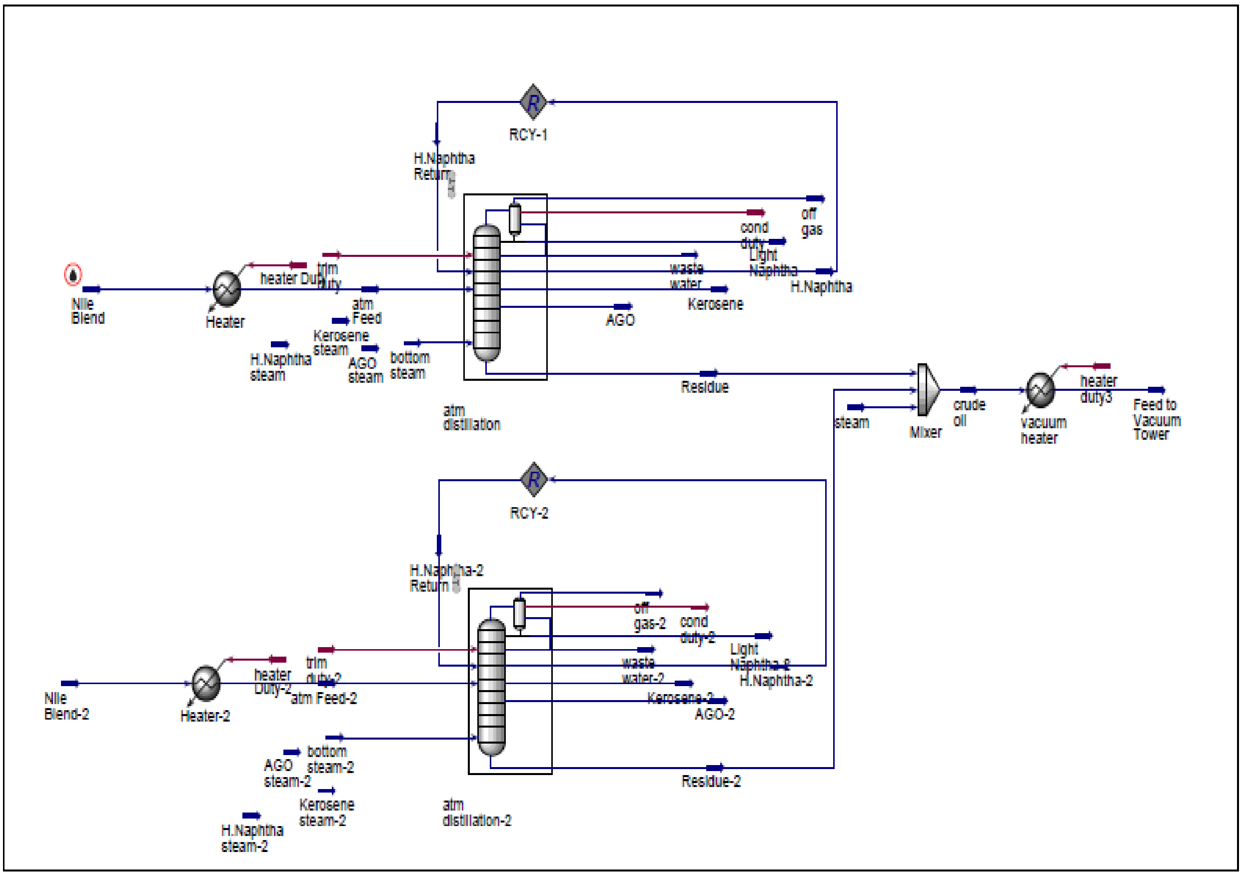
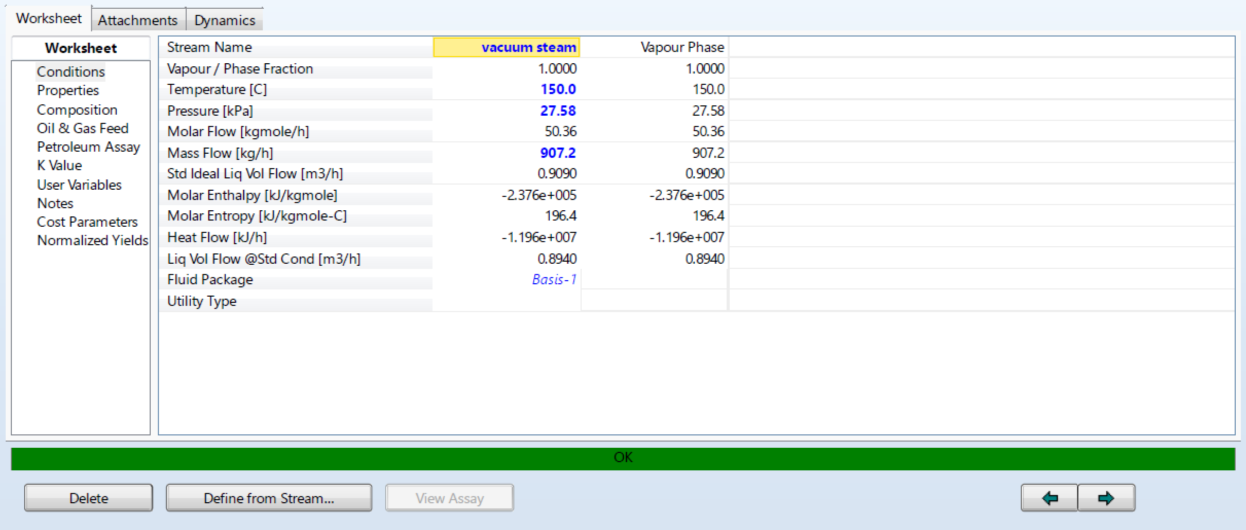


Figure3.62: The Heater on the flowsheet

A steam stream was added to the flowsheet, called (vacuum steam), and its specifications were entered as in (Figure 3.63) after that, in the same way as the previous tower addition, an absorption column was constructed on the flow sheet of the model palette (Figure. 3.64) and the basic information was filled out (Figure. 3.65).



Stream Name	vacuum steam	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	150.0	150.0
Pressure [kPa]	27.58	27.58
Molar Flow [kgmole/h]	50.36	50.36
Mass Flow [kg/h]	907.2	907.2
Std Ideal Liq Vol Flow [m3/h]	0.9090	0.9090
Molar Enthalpy [kJ/kgmole]	-2.376e+005	-2.376e+005
Molar Entropy [kJ/kgmole-C]	196.4	196.4
Heat Flow [kJ/h]	-1.196e+007	-1.196e+007
Liq Vol Flow @Std Cond [m3/h]	0.8940	0.8940
Fluid Package	Basis-1	
Utility Type		

Figure3.63: Definition of vacuum steam conditions

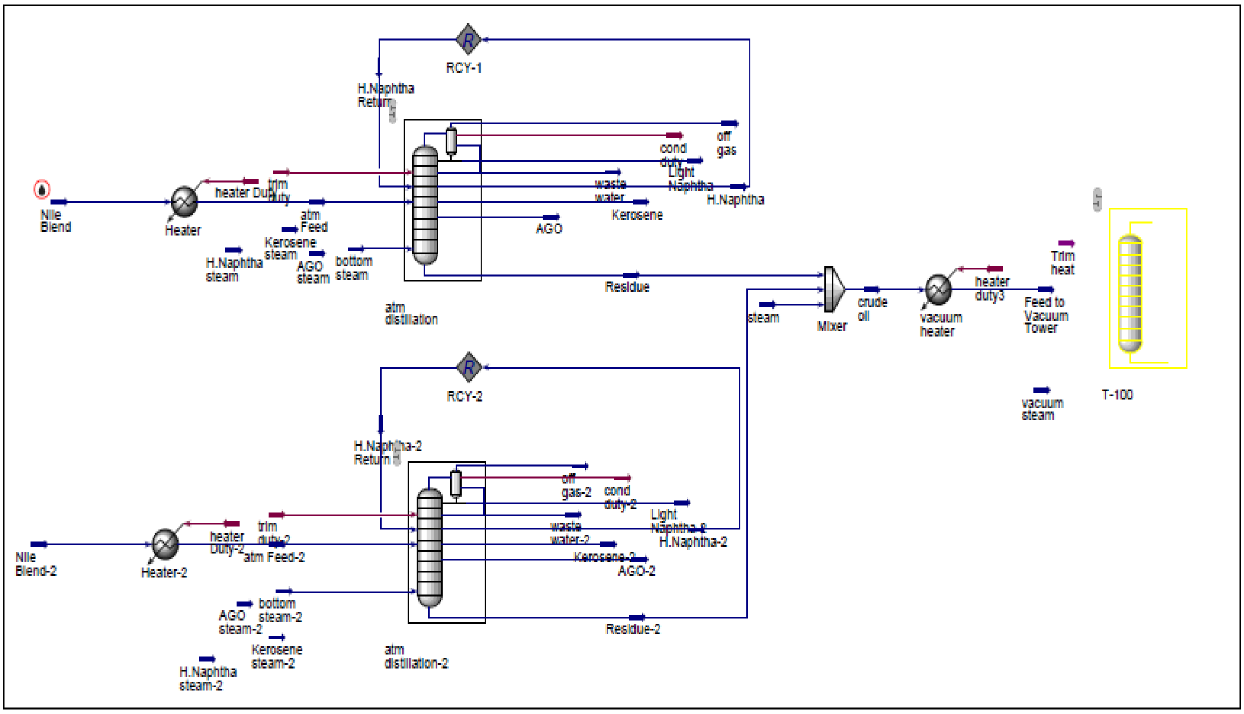
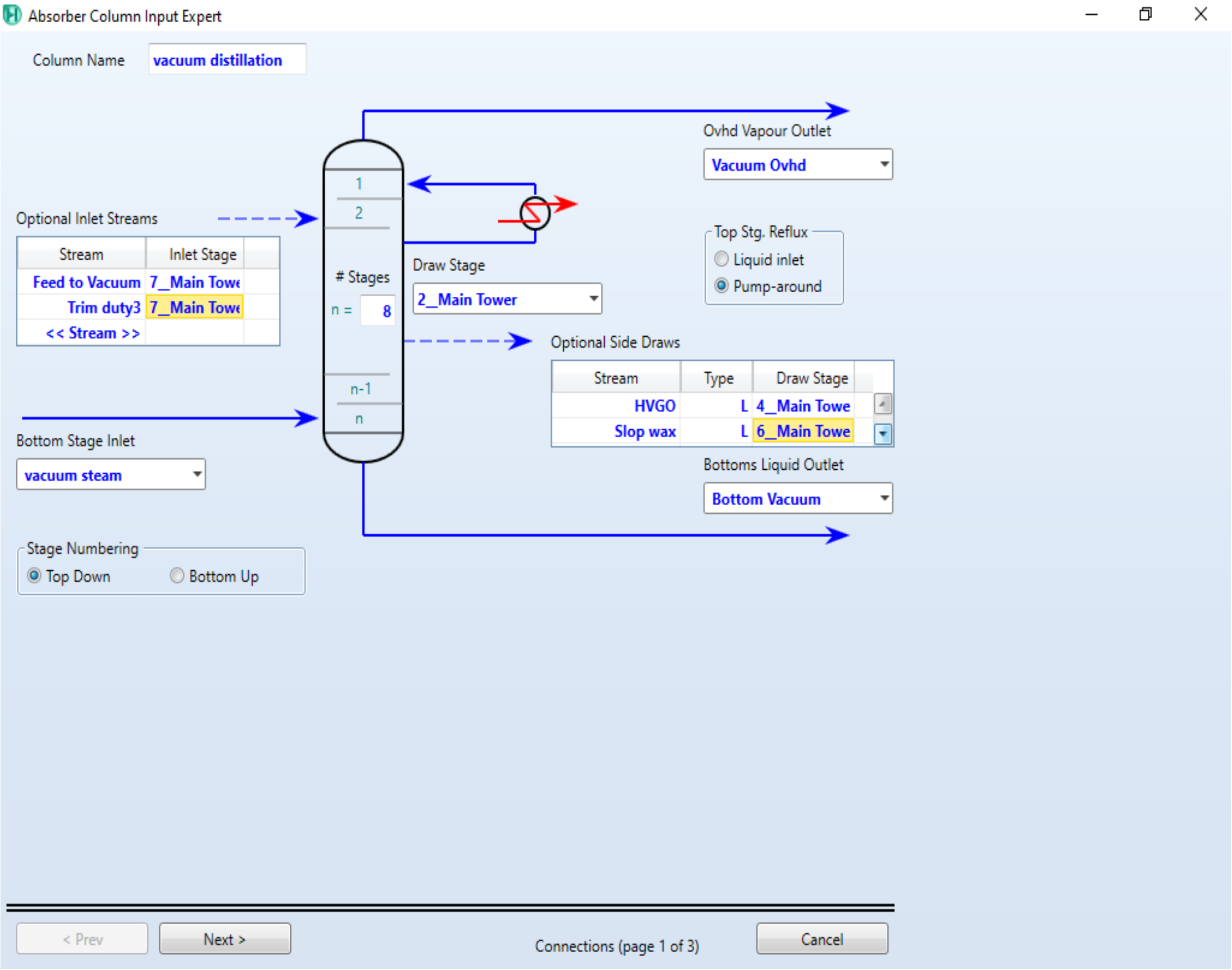


Figure3.64: Absorber column addition (vacuum distillation)



Column Name: vacuum distillation

Optional Inlet Streams

Stream	Inlet Stage
Feed to Vacuum	7__Main Towe
Trim duty3	7__Main Towe
<< Stream >>	

Bottom Stage Inlet: vacuum steam

Stage Numbering: ● Top Down ○ Bottom Up

Stages: n = 8

Draw Stage: 2__Main Tower

Ovhd Vapour Outlet: Vacuum Ovhd

Top Stg. Reflux: ○ Liquid inlet ● Pump-around

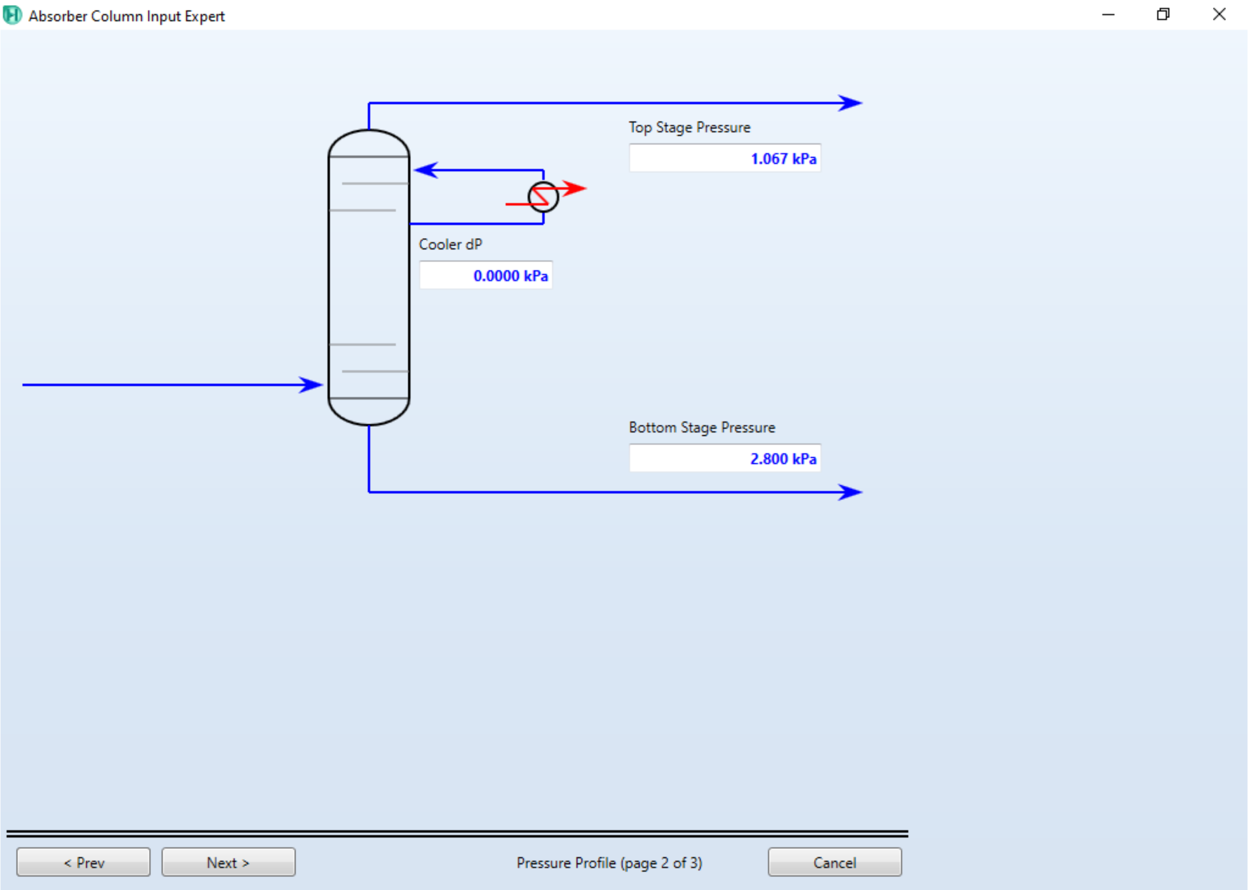
Optional Side Draws

Stream	Type	Draw Stage
HVGO	L	4__Main Towe
Slop wax	L	6__Main Towe

Bottoms Liquid Outlet: Bottom Vacuum

Figure3.65: The Basic Information of vacuum distillation column

The next button now becomes available, indicating sufficient information has been supplied to advance to the next page. Now click the next button to advance to the pressure profile page. On the pressure profile page, specify the top and bottom stage pressure of 1.067 kPa and 2.800 kPa respectively (Figure3.66).



Top Stage Pressure: 1.067 kPa

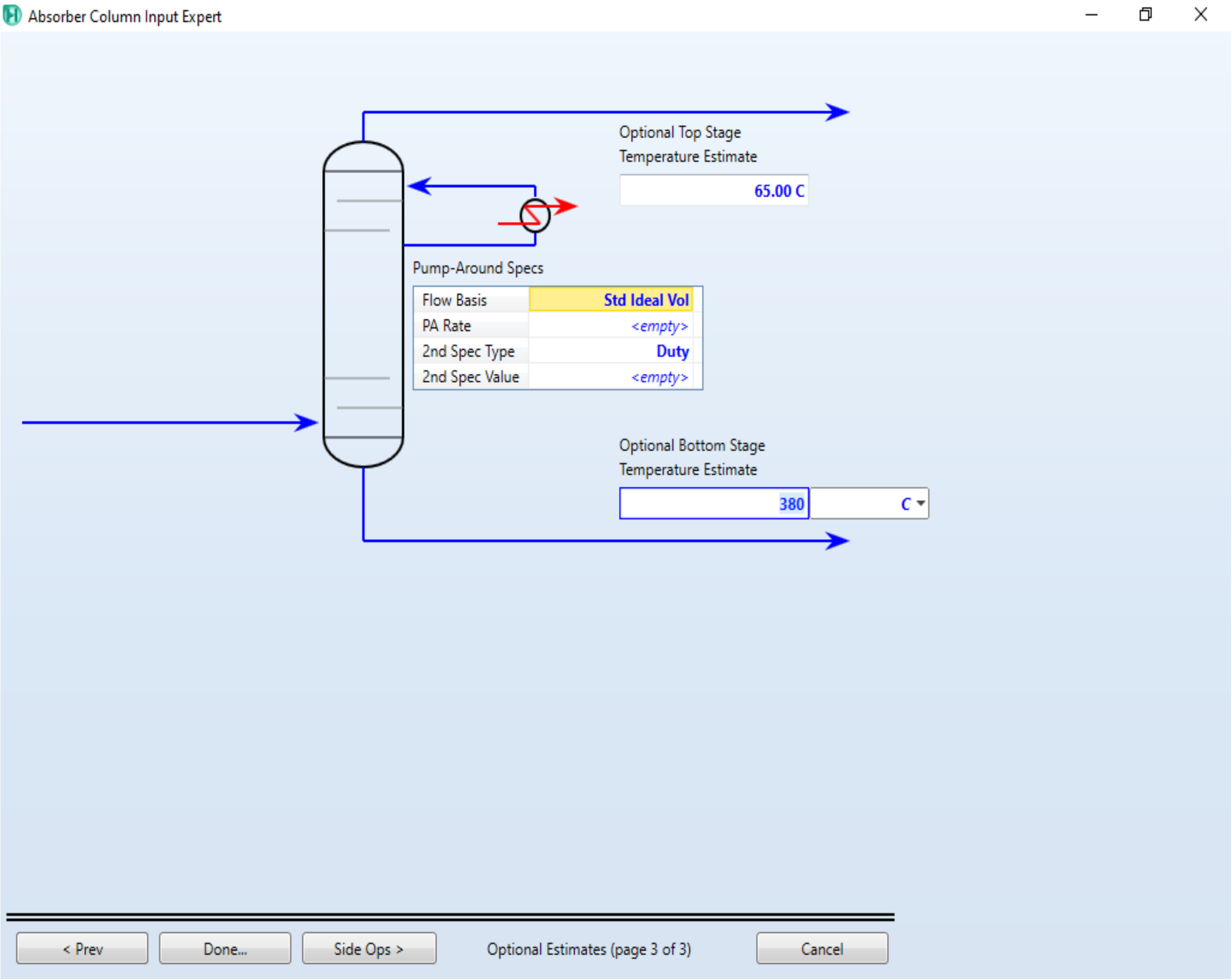
Cooler dP: 0.0000 kPa

Bottom Stage Pressure: 2.800 kPa

Pressure Profile (page 2 of 3)

Figure3.66: Pressure profile (vacuum distillation)

Click the next button to advance to the optional estimates page. Although HYSYS does not usually require estimates to produce a converged column, the good estimates result in a faster solution. Specify the temperature of top stage 65.00˚C and temperature of bottom stage 380.00˚C (Figure3.67).



Optional Top Stage Temperature Estimate: 65.00 C

Pump-Around Specs

Flow Basis	Std Ideal Vol
PA Rate	<empty>
2nd Spec Type	Duty
2nd Spec Value	<empty>

Optional Bottom Stage Temperature Estimate: 380 C

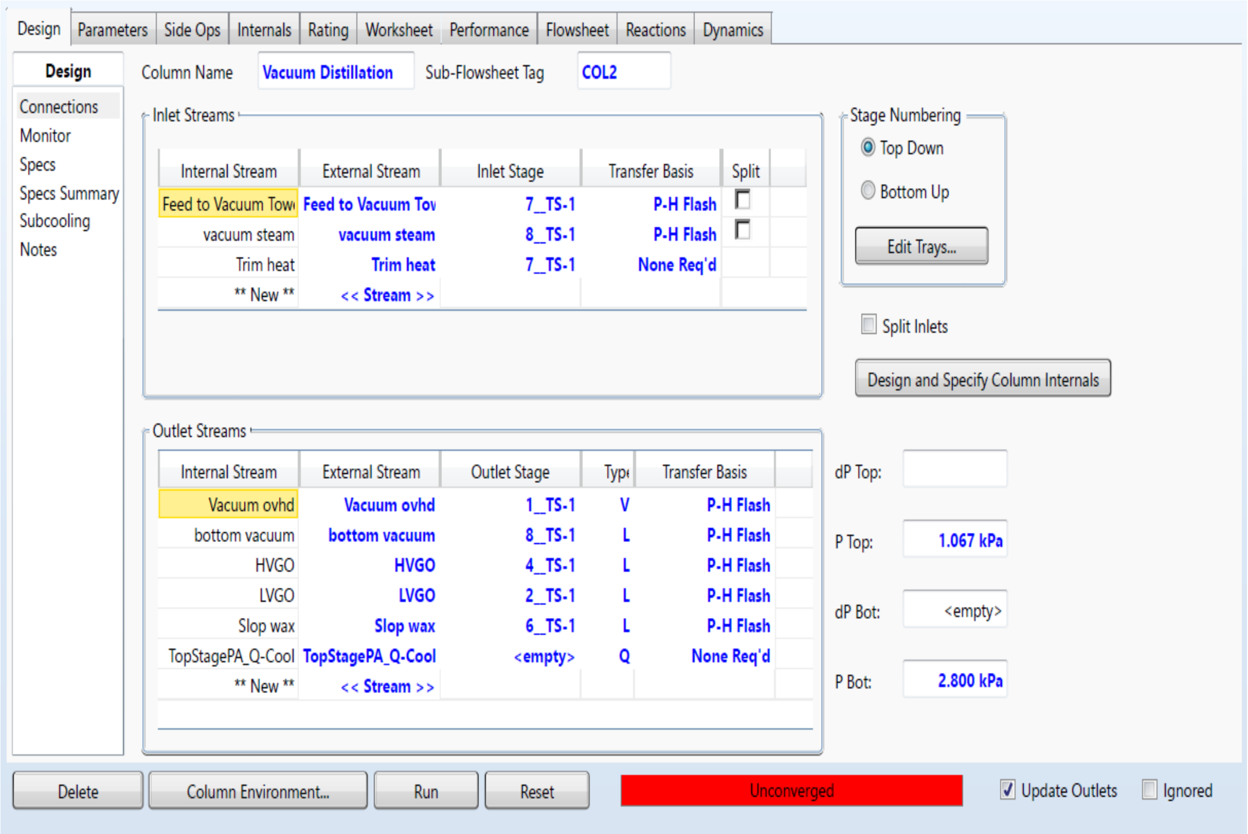
Optional Estimates (page 3 of 3)

Figure3.67: Optional Estimates

The current degrees of freedom are equal to 1 degree (Figure 3.69), so add the temperature specification on top of the tower (ovhd temperature) and then the following specifications were added (tray 8 temperature, vacuum ovhd rate, bottom rate, and Top stage PA-T Ret(Pa)) as shown in the (Figure 3.70).

After that deactivate LVGO rate, Top stage PA-Duty (Pa) and activate vacuum ovhd rate, Top stage PA-T Ret(Pa) and tray 8 temperature.

Select the screen page to view the specification matrix. Again, degrees of freedom are equal to zero, so the column is ready to calculate the flow rate of the vacuum distillate and we make final modifications to the specification to obtain convergence (Figure 3.71).



Column Name: Vacuum Distillation
Sub-Flowsheet Tag: COL2

Inlet Streams

Internal Stream	External Stream	Inlet Stage	Transfer Basis	Split
Feed to Vacuum Tow	Feed to Vacuum Tov	7__TS-1	P-H Flash	☐
vacuum steam	vacuum steam	8__TS-1	P-H Flash	☐
Trim heat	Trim heat	7__TS-1	None Req'd	
** New **	<< Stream >>			

Stage Numbering
● Top Down
○ Bottom Up

☐ Split Inlets

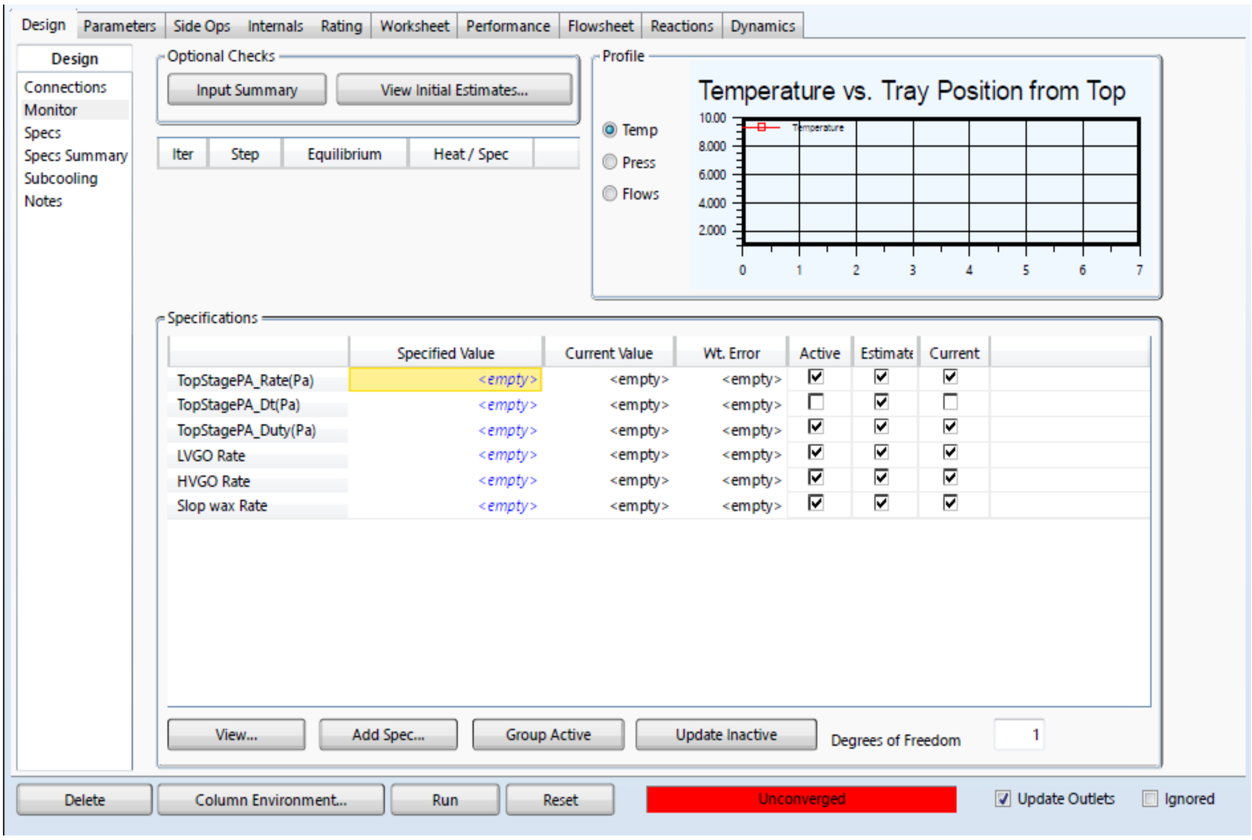
Outlet Streams

Internal Stream	External Stream	Outlet Stage	Typ	Transfer Basis
Vacuum ovhd	Vacuum ovhd	1__TS-1	V	P-H Flash
bottom vacuum	bottom vacuum	8__TS-1	L	P-H Flash
HVGO	HVGO	4__TS-1	L	P-H Flash
LVGO	LVGO	2__TS-1	L	P-H Flash
Slop wax	Slop wax	6__TS-1	L	P-H Flash
TopStagePA_Q-Cool	TopStagePA_Q-Cool	<empty>	Q	None Req'd
** New **	<< Stream >>			

dP Top:
P Top: 1.067 kPa
dP Bot: <empty>
P Bot: 2.800 kPa

Unconverged
☑ Update Outlets
☐ Ignored

Figure3.68: Connections of inlet and outlet streams from design



Profile: ● Temp ○ Press ○ Flows

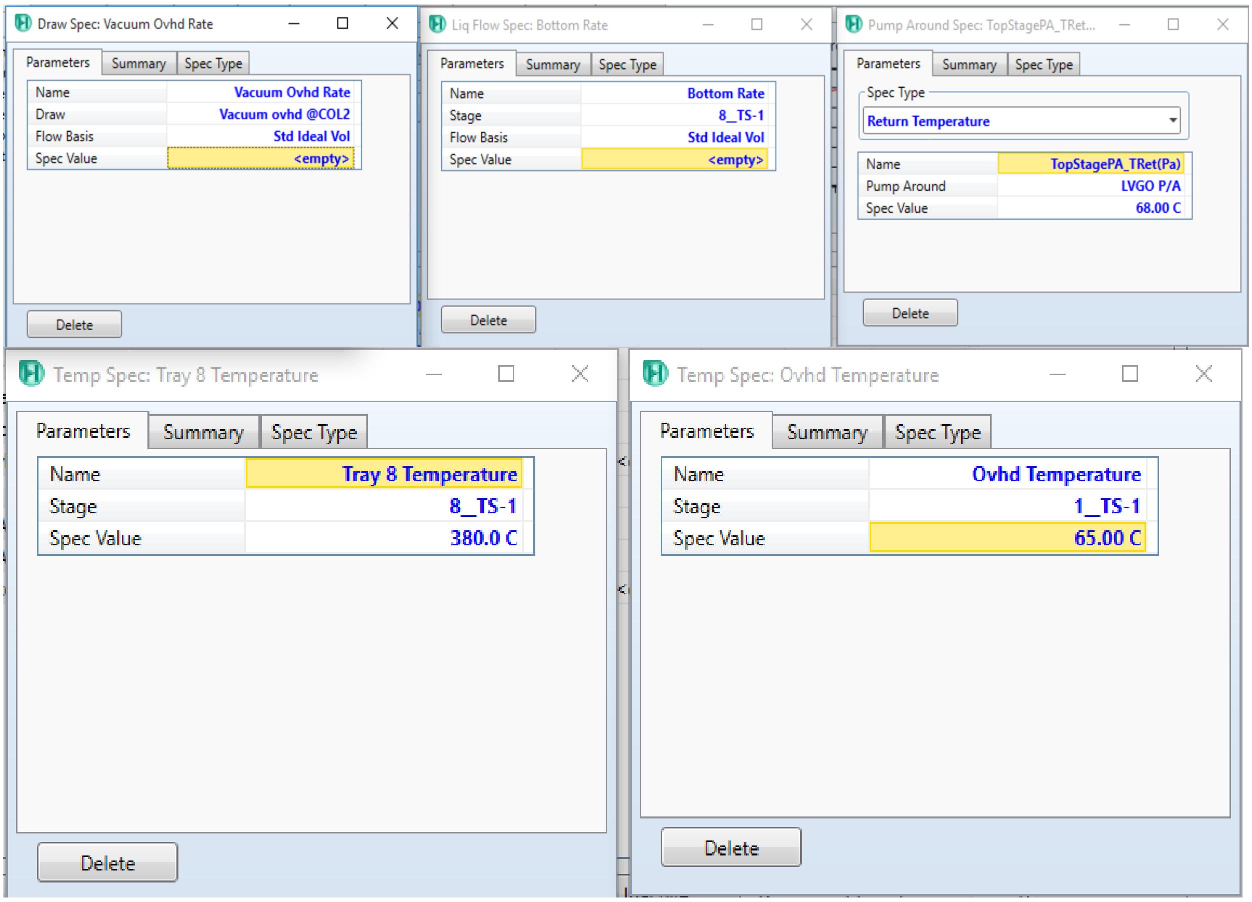
Specifications

	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
TopStagePA_Rate(Pa)	<empty>	<empty>	<empty>	☑	☑	☑
TopStagePA_Dt(Pa)	<empty>	<empty>	<empty>	☐	☑	☐
TopStagePA_Duty(Pa)	<empty>	<empty>	<empty>	☑	☑	☑
LVGO Rate	<empty>	<empty>	<empty>	☑	☑	☑
HVGO Rate	<empty>	<empty>	<empty>	☑	☑	☑
Slop wax Rate	<empty>	<empty>	<empty>	☑	☑	☑

Degrees of Freedom: 1

Unconverged
☑ Update Outlets
☐ Ignored

Figure3.69: Degrees of freedom (vacuum distillation)



Name	Vacuum Ovhd Rate
Draw	Vacuum ovhd @COL2
Flow Basis	Std Ideal Vol
Spec Value	<empty>

Name	Bottom Rate
Stage	8_TS-1
Flow Basis	Std Ideal Vol
Spec Value	<empty>

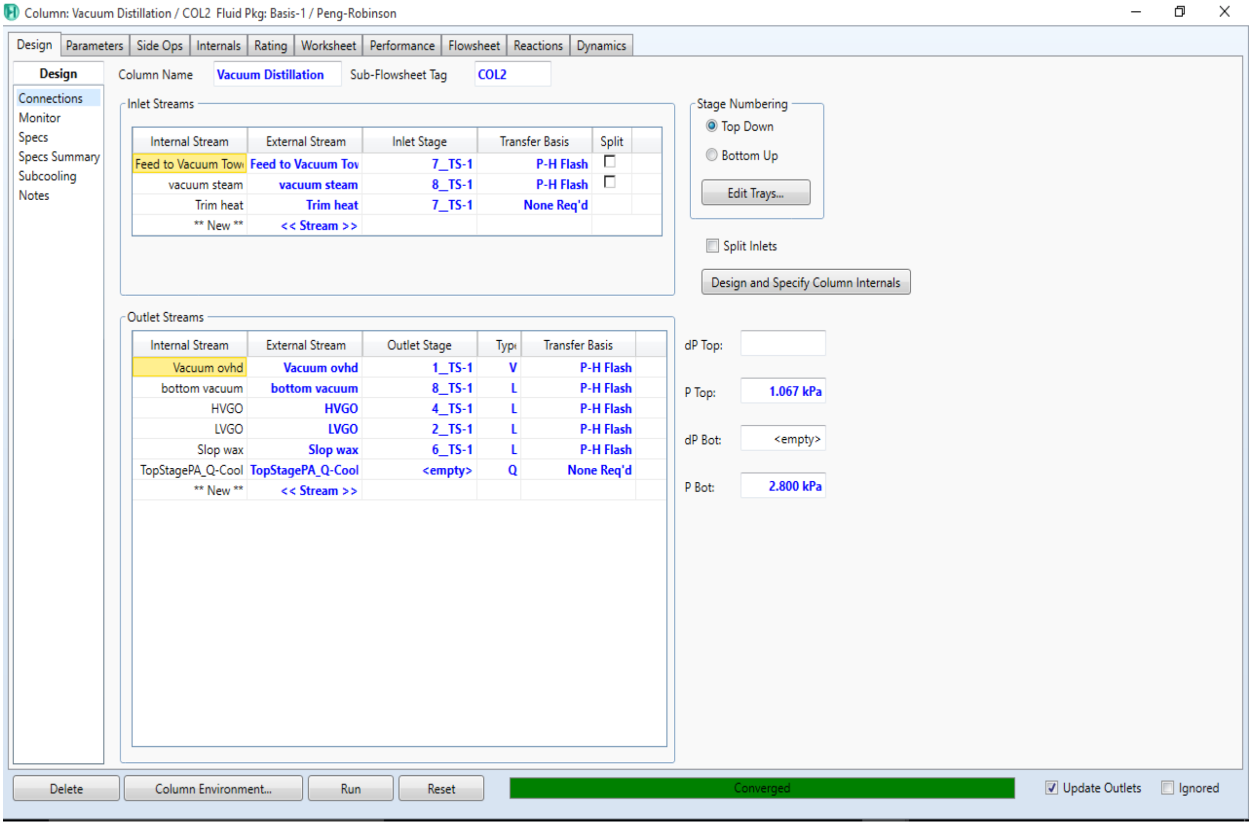
Spec Type: Return Temperature

Name	TopStagePA_TRet(Pa)
Pump Around	LVGO P/A
Spec Value	68.00 C

Name	Tray 8 Temperature
Stage	8_TS-1
Spec Value	380.0 C

Name	Ovhd Temperature
Stage	1_TS-1
Spec Value	65.00 C

Figure3.70: Specifications addition to complete the definition



Column Name: Vacuum Distillation
Sub-Flowsheet Tag: COL2

Inlet Streams

Internal Stream	External Stream	Inlet Stage	Transfer Basis	Split
Feed to Vacuum Tow	Feed to Vacuum Tov	7_TS-1	P-H Flash	☐
vacuum steam	vacuum steam	8_TS-1	P-H Flash	☐
Trim heat	Trim heat	7_TS-1	None Req'd	
** New **	<< Stream >>			

Stage Numbering
● Top Down
○ Bottom Up

☐ Split Inlets

Outlet Streams

Internal Stream	External Stream	Outlet Stage	Typ	Transfer Basis
Vacuum ovhd	Vacuum ovhd	1_TS-1	V	P-H Flash
bottom vacuum	bottom vacuum	8_TS-1	L	P-H Flash
HVGO	HVGO	4_TS-1	L	P-H Flash
LVGO	LVGO	2_TS-1	L	P-H Flash
Slop wax	Slop wax	6_TS-1	L	P-H Flash
TopStagePA_Q-Cool	TopStagePA_Q-Cool	<empty>	Q	None Req'd
** New **	<< Stream >>			

dP Top:
P Top: 1.067 kPa
dP Bot: <empty>
P Bot: 2.800 kPa

☑ Update Outlets ☐ Ignored

Figure3.71: Running the vacuum distillation tower

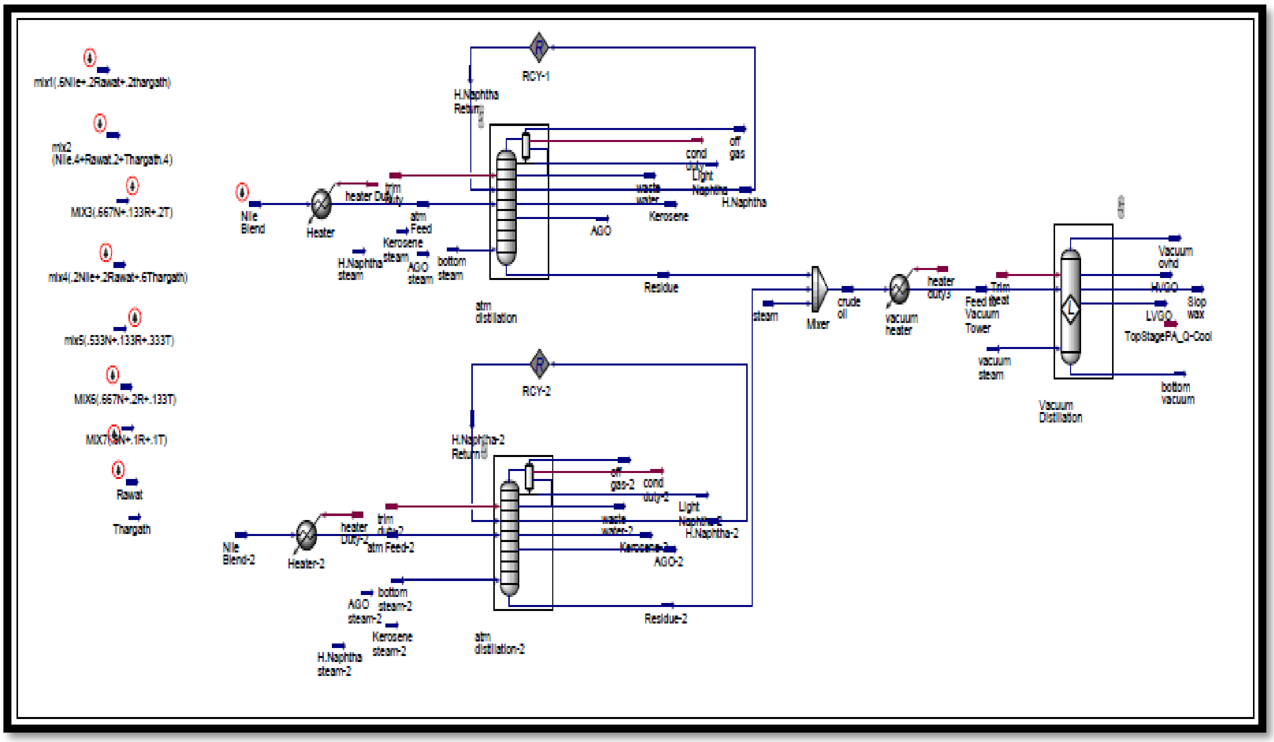


Figure3.72: The proposed new shape of El Obeid refinery

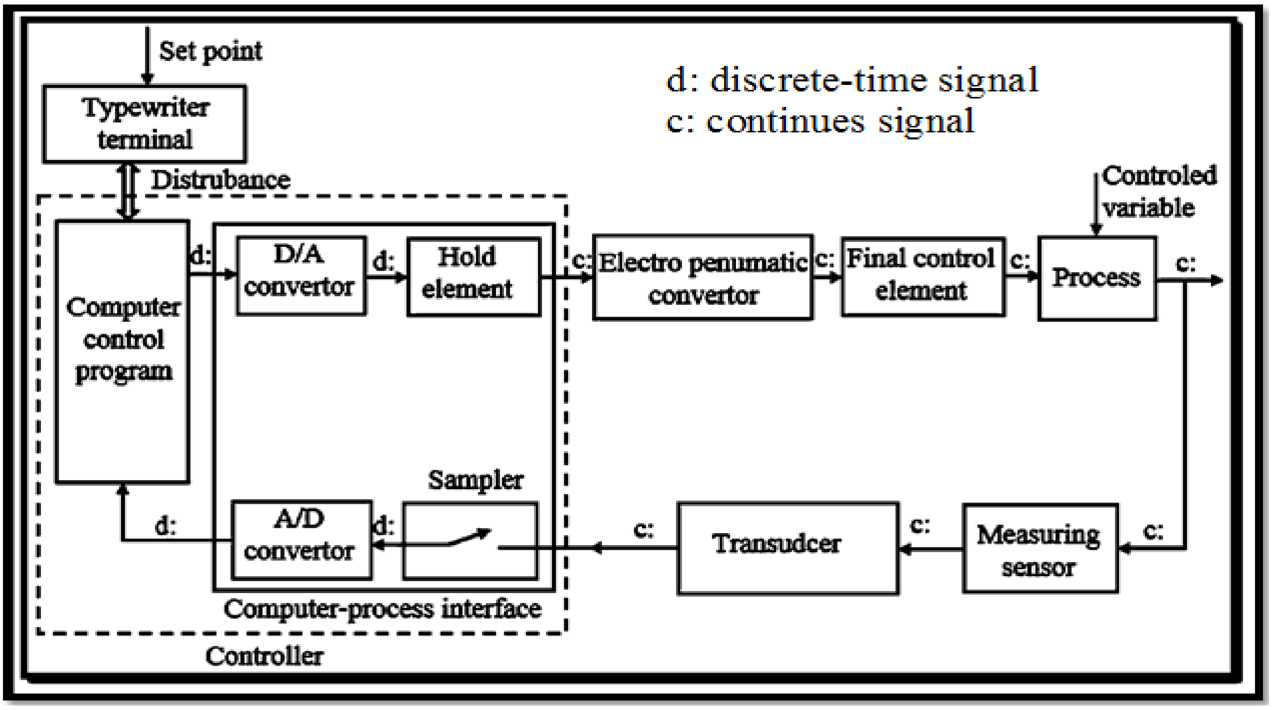


Figure 3.73: The digital control system

CHAPTER FOUR

Results and Discussion

4.1 Introduction

This chapter presents the results and discussion of the various parts of the work that were conducted during the execution of this study. These include crude oil characterization, simulation of PDF, assumption validation, retrofit measures.

4.2 Properties of Crude oil

The crude oil was characterized using the true boiling point (TBP) analysis methodology. Observation to the variation in all crude oil properties, this can be attributed to the variation on reservoir, and any small differences in composition can greatly affect the physical properties and processing required for producing salable products.

The Crude oils properties was analysis of (Nile Blend, Rawat, Thargath, Mix 1, Mix 2, Mix 3, Mix 4, Mix 5, Mix 6 and Mix7) using Hysys software summarized in Table 4.1.

4.3 Process Description

Concerning the aforementioned subject, the previous Crude oils were studied according to simulation program {HYSYS} to find out if EL Obeid Refinery Company is capable to refine it and produce the desired values. All of our operation conditions and equipment design data, was taken into consideration.

The first step of study, prepare the feed of 15,000 bbl/day of pure (Nile Blend, Rawat and Thargath), and using Hysys for modifications needed to facilitate, heating and handling of the crude in the software before applying it in practice.

So, next step to make alternatives by blending (Nile Blend, Rawat and Thargath) to get an optimum ratio that can be handled by ORC plant without major modifications or high risk of processing.

Table 4.1: Properties of Different crudes

Properties	Nile Blend	Rawat	Thargath	Mix 1	Mix2	Mix3	Mix4	Mix5	Mix6	Mix 7
Molecular Weight	326.8	317.8	443	343.5	363.9	344.1	386.5	357.5	337.1	335
Mass Density (API)	27.78	30.74	22.35	27.24	26.13	27.05	25.04	26.31	27.61	27.51
Mass Exergy (kg/kJ)	0.1317	0.1312	0.131	0.1314	0.1313	0.1314	0.1311	0.1313	0.1314	0.1315
Specific Heat (Kj/kgmole.C)	619.3	600	835.2	649.5	687.4	650.8	729.6	675.8	637.7	633.8
Std.Ideal liq.Mass Density (API-60)	25.74	28.62	20.51	25.22	24.16	25.04	23.1	24.32	25.58	25.48
Watson K	11.79	11.88	11.9	11.82	11.84	11.81	11.86	11.82	11.82	11.8
C_P/C_V	1.131	1.157	1.154	1.134	1.134	1.133	1.14	1.134	1.134	1.131
Kinematic Viscosity (cSt)	29.85	25.26	124.9	35.1	44.69	35.73	58.49	41.93	32.54	32.18
Surface Tension (dyne/cm)	26.82	27.58	29.35	27.4	27.86	27.35	28.38	27.65	27.26	27.11
Thermal Conductivity (W/m.k)	0.1238	0.1246	0.1301	0.125	0.1261	0.1249	0.1273	0.1256	0.1246	0.1244
Bubble Point Pressure (k pa)	7.478	7.191	1.234	6.517	5.539	6.531	4.393	5.911	6.814	6.982
Viscosity (cp)	26.46	21.99	114.7	31.23	40.04	31.82	52.77	37.52	28.88	28.85
Viscosity Index	28.86	28.15	33.88	29.51	30.45	29.59	31.42	30.2	29.21	29.17

4.4 Atmospheric Distillation Tower Results

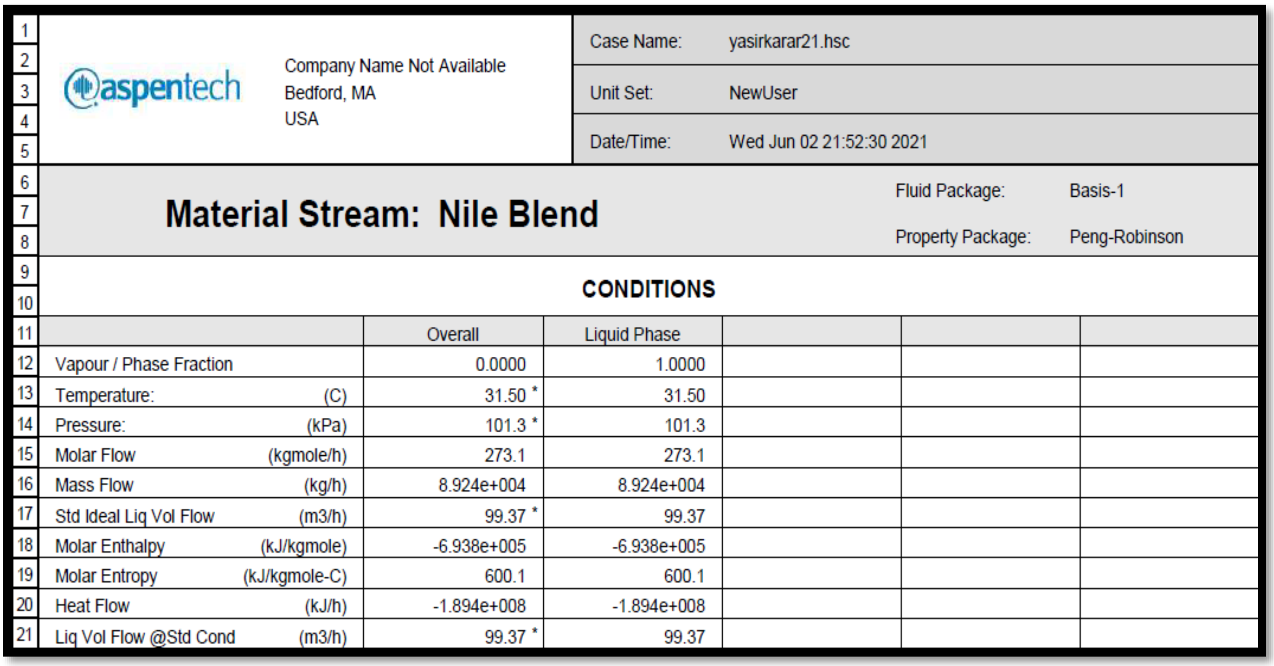
4.4.1 Case Study no.1 (Blend 1)

Pure Nile Blend crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.2: The Oil Distributions-Blend 1

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	34.4	59	0.015
Light Naphtha	59	175	0.079
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.084
AGO	260	380	0.180
Residue	380	1014	0.637

(Figure 4.1) showing the Conditions of pure Nile Blend crude before entering the main distillation column.



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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 21:52:30 2021

Material Stream: Nile Blend

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (C)	31.50 *	31.50
Pressure: (kPa)	101.3 *	101.3
Molar Flow (kgmole/h)	273.1	273.1
Mass Flow (kg/h)	8.924e+004	8.924e+004
Std Ideal Liq Vol Flow (m3/h)	99.37 *	99.37
Molar Enthalpy (kJ/kgmole)	-6.938e+005	-6.938e+005
Molar Entropy (kJ/kgmole-C)	600.1	600.1
Heat Flow (kJ/h)	-1.894e+008	-1.894e+008
Liq Vol Flow @Std Cond (m3/h)	99.37 *	99.37

Figure 4.1: Conditions of Nile blend stream (Blend 1)

The following figures showing TBP curve (Figure 4.2) and Cuts Distribution (Figure 4.3).

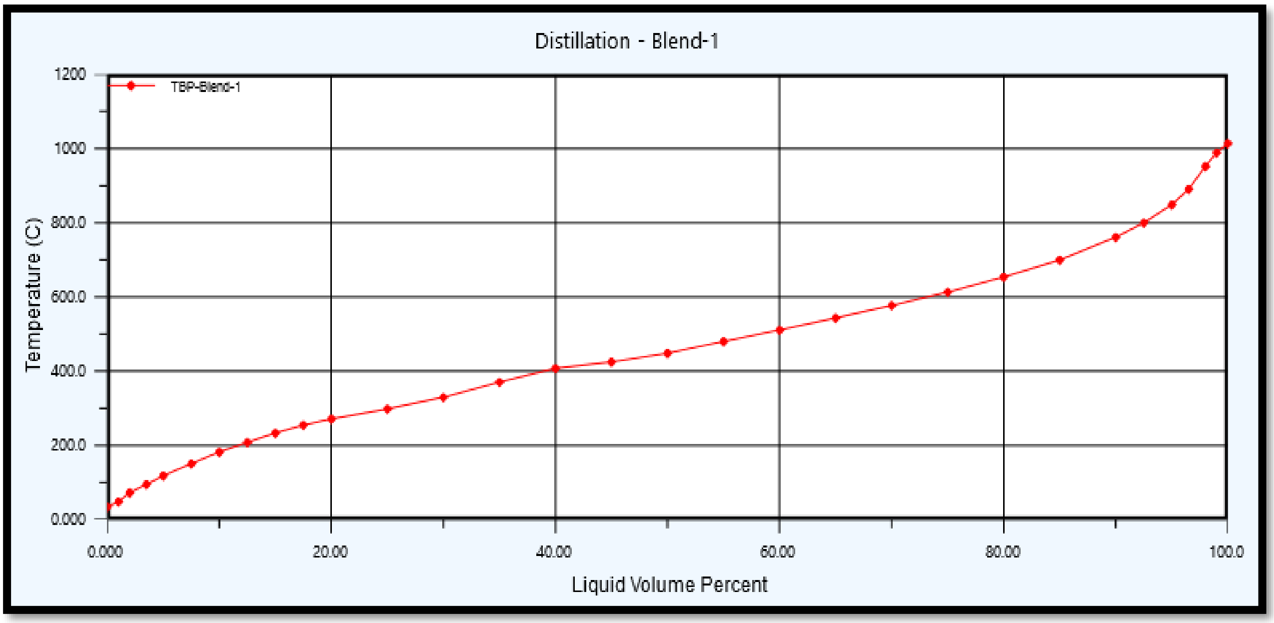


Figure 4.2: True Boiling Point Distillation Curve-Blend 1

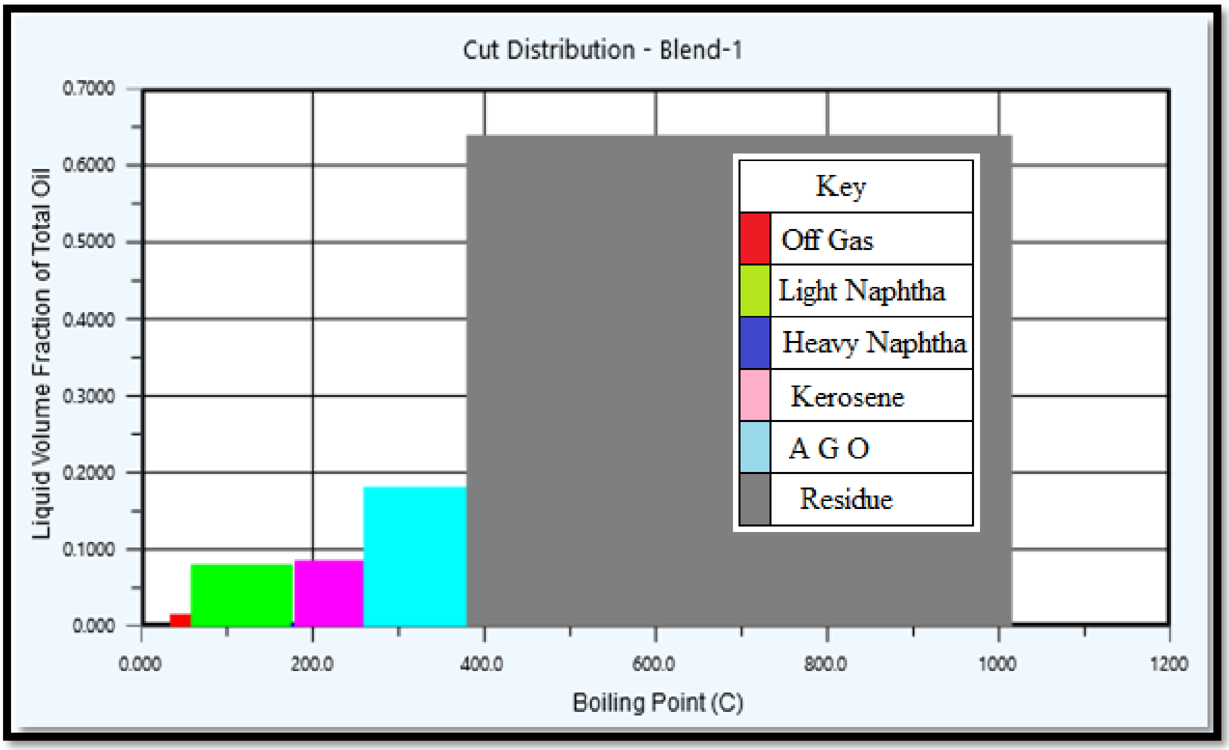


Figure 4.3: Cut Distribution-Blend 1

(Figure 4.4) showing the increasing of pressure during process

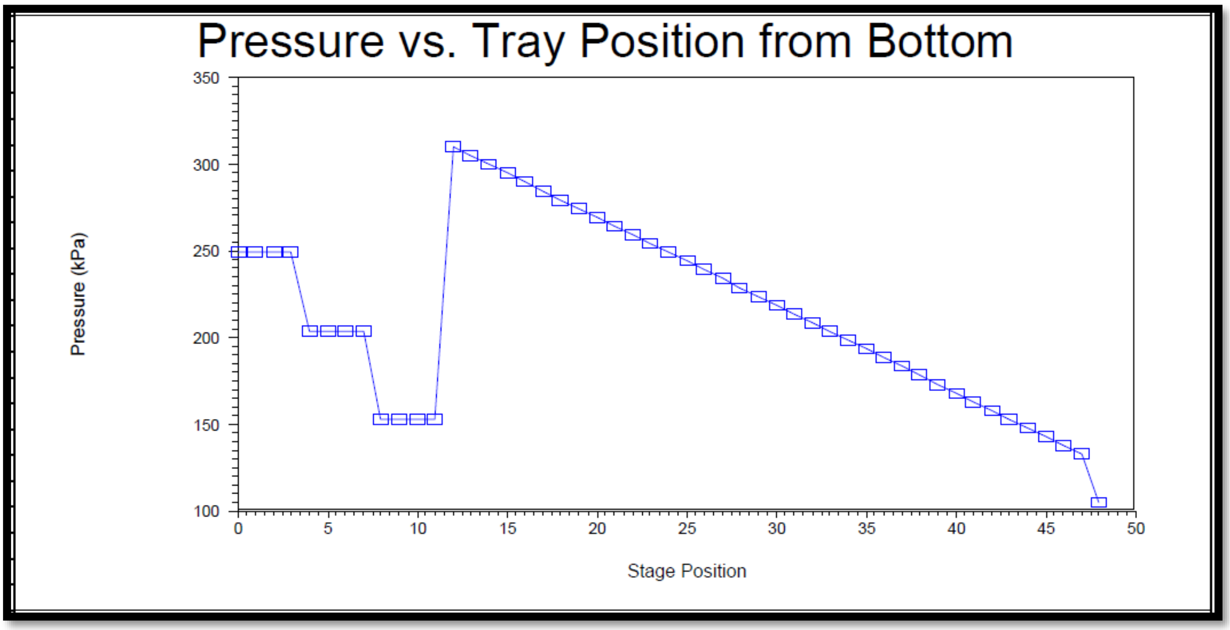


Figure4.4: Pressure profile (Blend 1)

The change of temperature from tray to another shown in (figure 4.5)

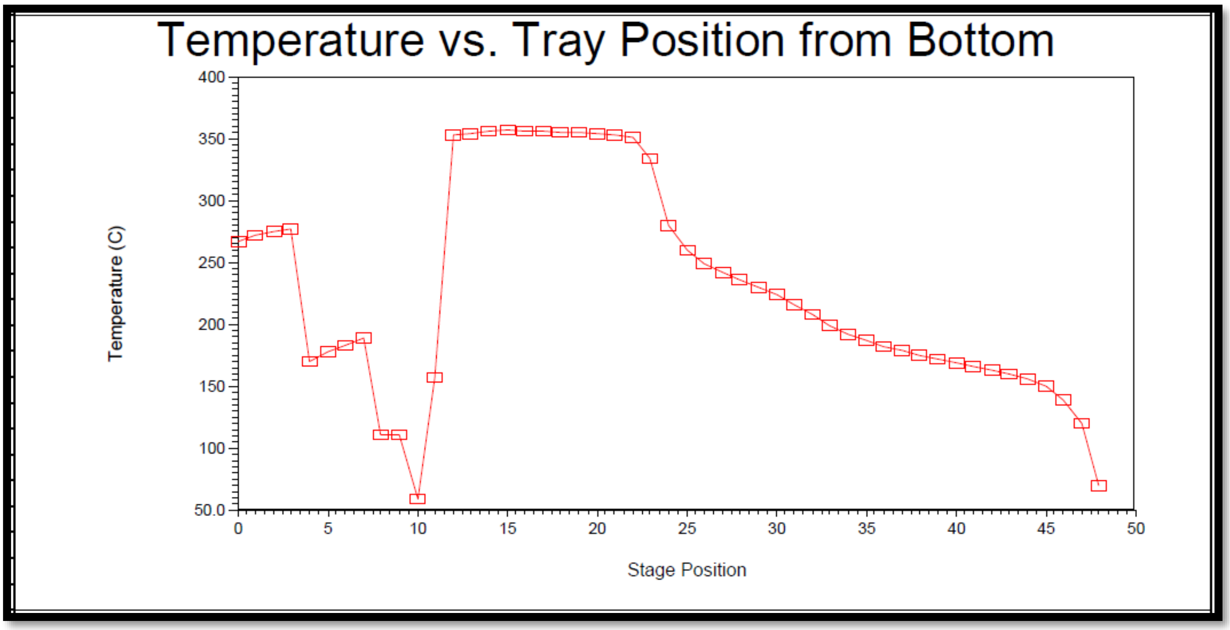


Figure 4.5: Temperature profile (Blend 1)

The quantity of liquid flow and Vapour flow shown in (figure 4.6)

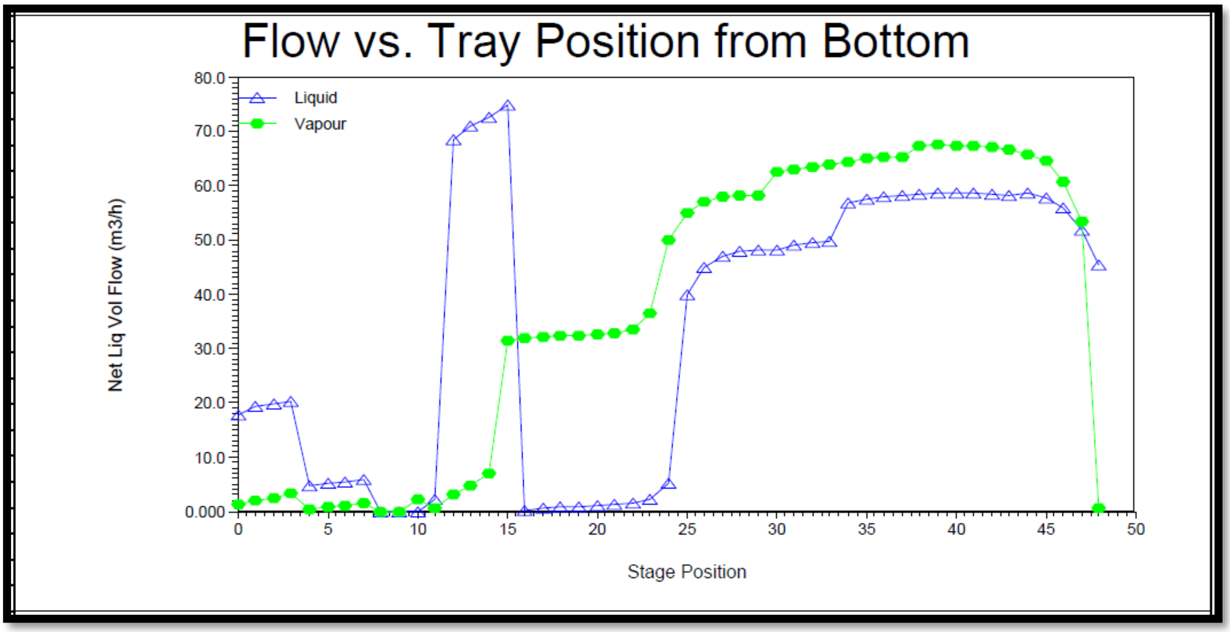


Figure 4.6: Net flow profile (Blend 1)

Column properties (Density and Molecular weight) shown in (figure 4.7).

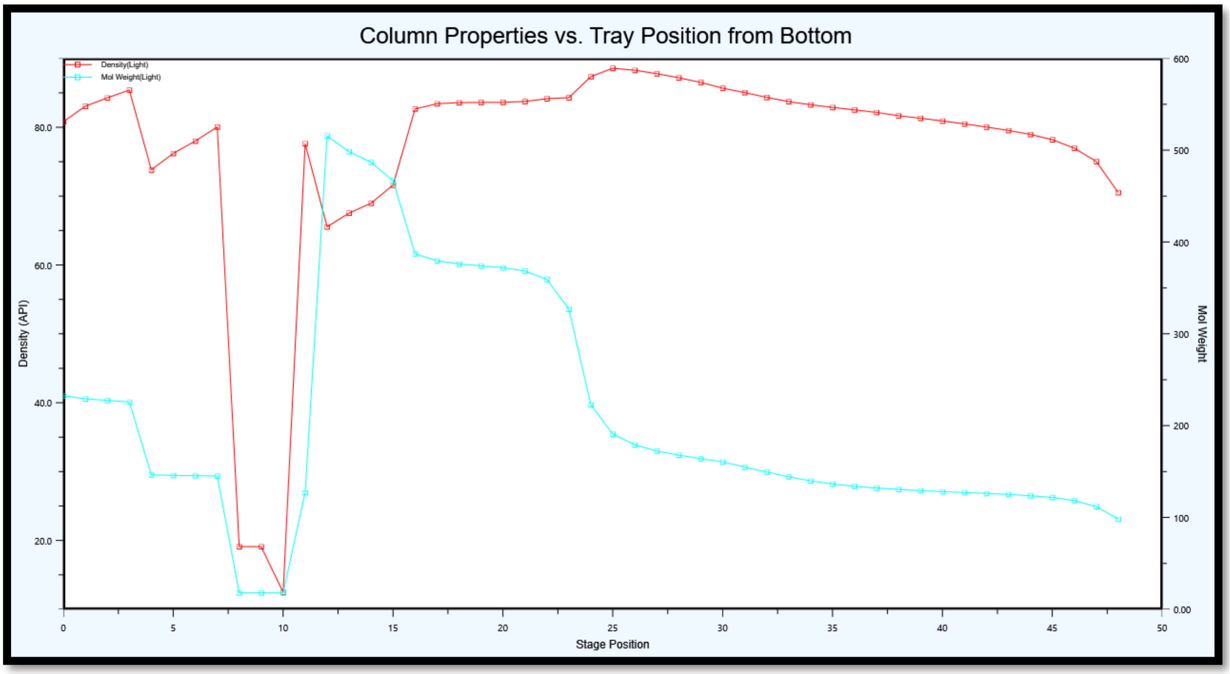


Figure 4.7: Correlation between density and molecular weight (Blend 1)

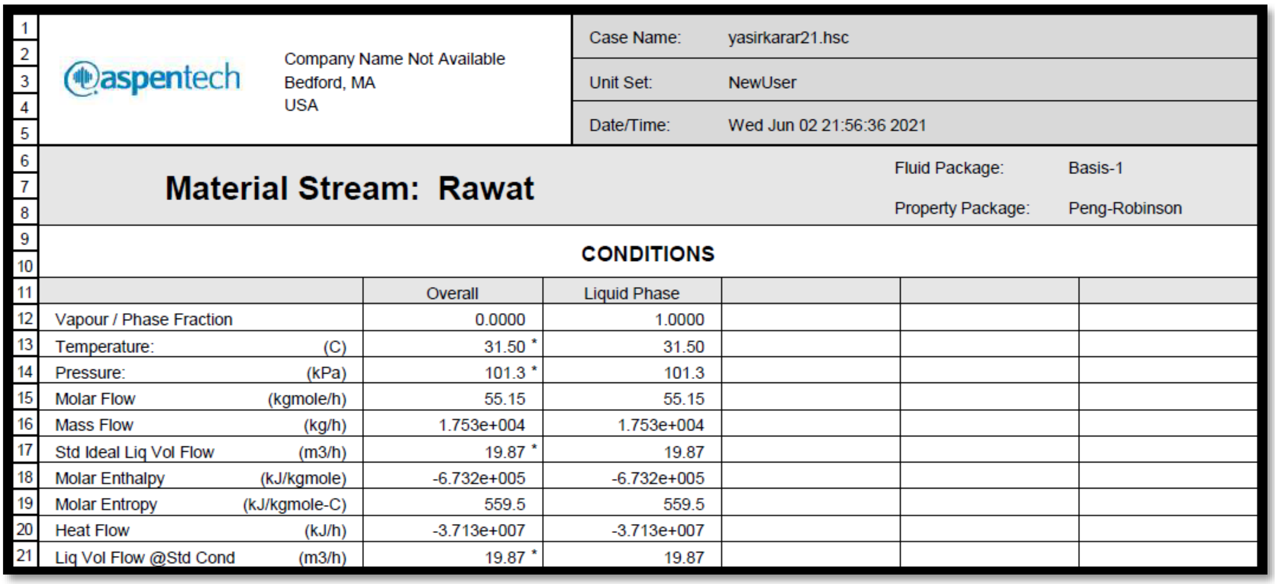
4.4.2 Case Study no.2 (Blend 2)

Pure Rawat crude was chosen, displaying the TBP curve and cutting the distribution and the following results are obtained from Hysys.

Table 4.3: The Oil Distributions -Blend 2

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	33.84	59	0.017
Light Naphtha	59	175	0.049
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.078
AGO	260	380	0.252
Residue	380	564.3	0.601

(Figure 4.8) showing the Conditions of pure Rawat crude before entering the main distillation column.



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Unit Set: NewUser
Date/Time: Wed Jun 02 21:56:36 2021

Material Stream: Rawat

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

		Overall	Liquid Phase
Vapour / Phase Fraction		0.0000	1.0000
Temperature:	(C)	31.50 *	31.50
Pressure:	(kPa)	101.3 *	101.3
Molar Flow	(kgmole/h)	55.15	55.15
Mass Flow	(kg/h)	1.753e+004	1.753e+004
Std Ideal Liq Vol Flow	(m3/h)	19.87 *	19.87
Molar Enthalpy	(kJ/kgmole)	-6.732e+005	-6.732e+005
Molar Entropy	(kJ/kgmole-C)	559.5	559.5
Heat Flow	(kJ/h)	-3.713e+007	-3.713e+007
Liq Vol Flow @Std Cond	(m3/h)	19.87 *	19.87

Figure 4.8: Conditions of Rawat stream (Blend 2)

The following figures showing TBP curve (Figure 4.9) and Cuts Distribution (Figure 4.10).

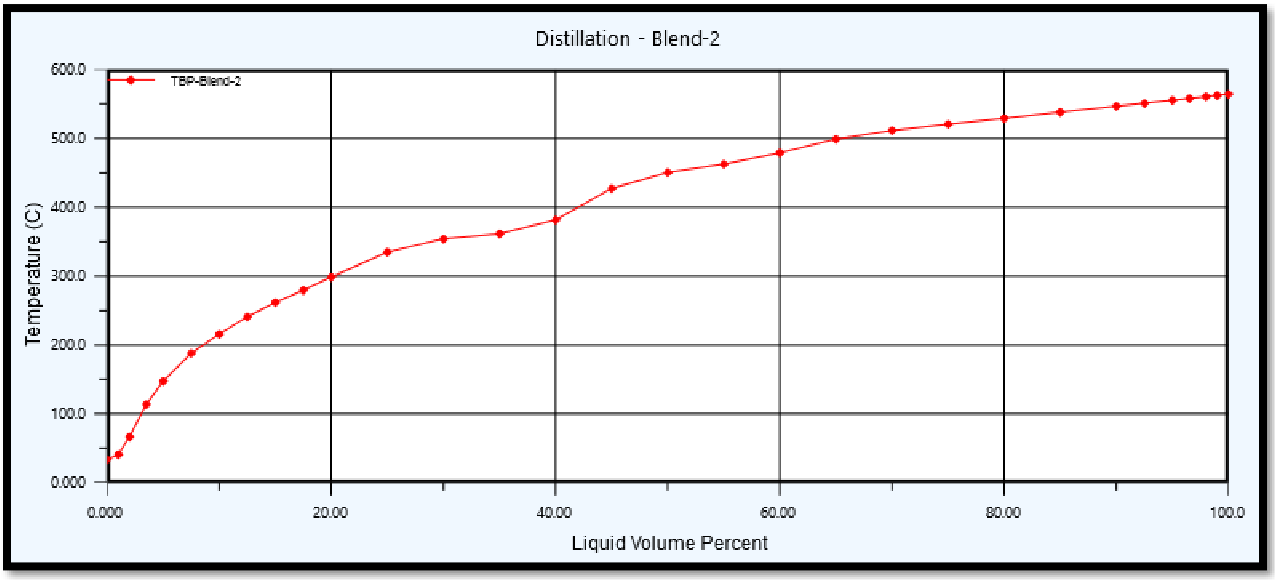


Figure 4.9: True Boiling Point Distillation Curve-Blend 2

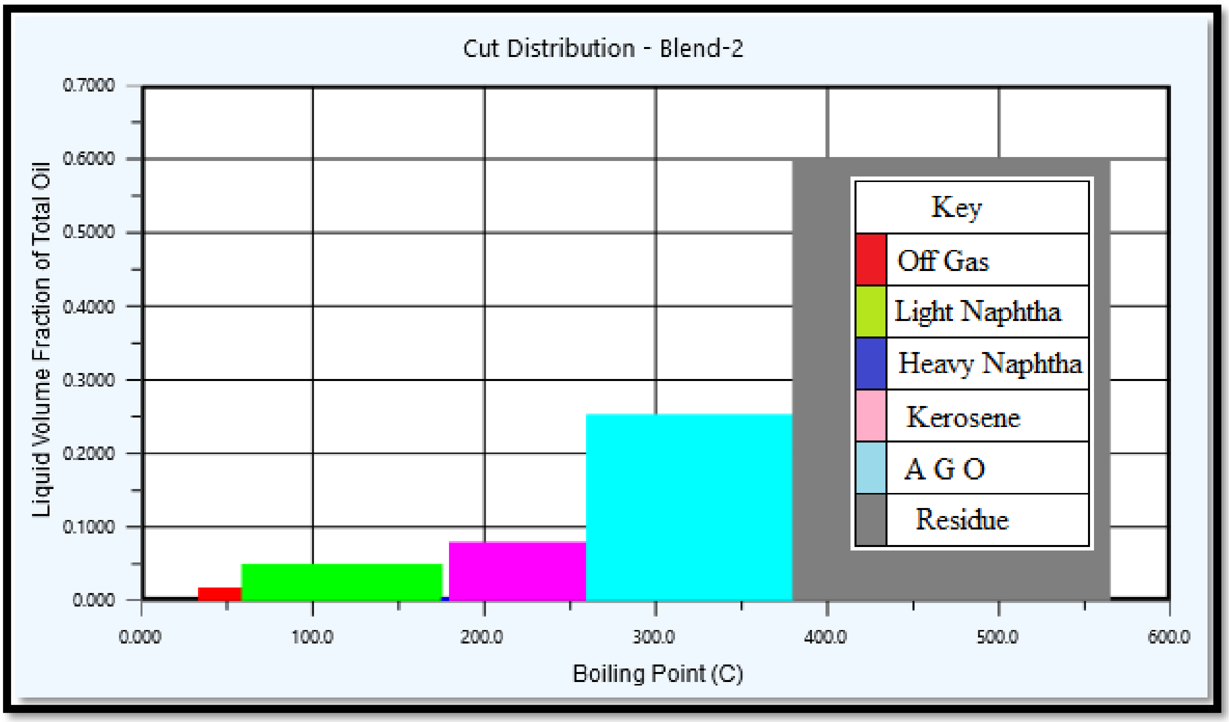


Figure 4.10: Cut Distribution-Blend 2

(Figure 4.11) showing the increasing of pressure during process

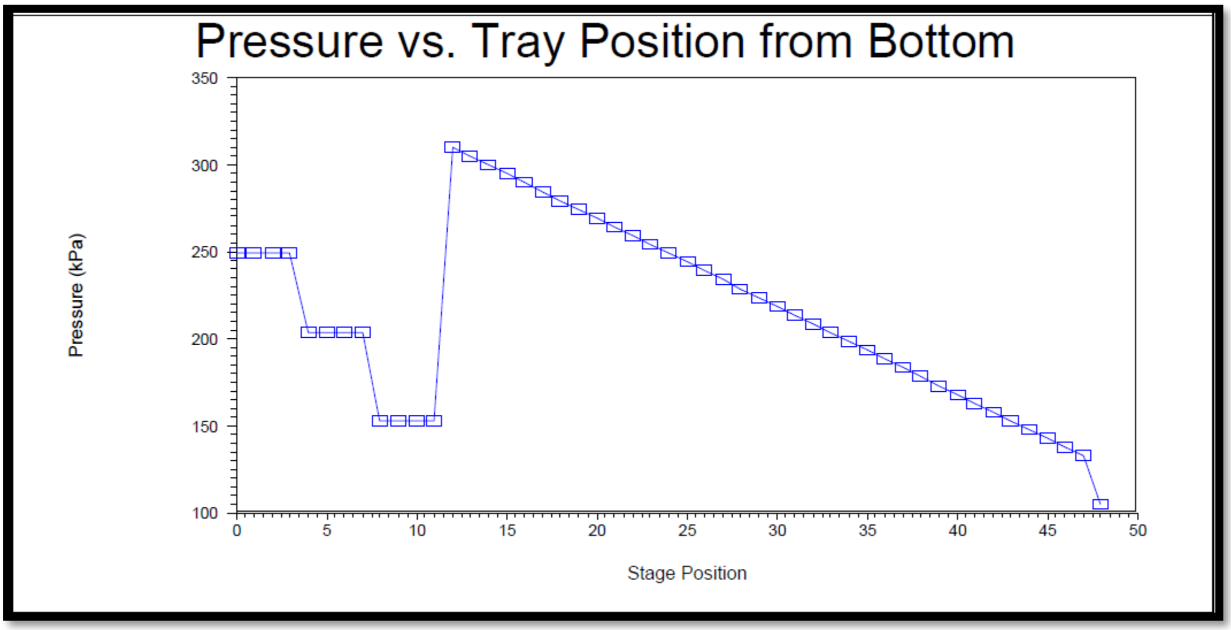


Figure 4.11: Pressure profile (Blend 2)

The change of temperature from tray to another shown in (figure 4.12)

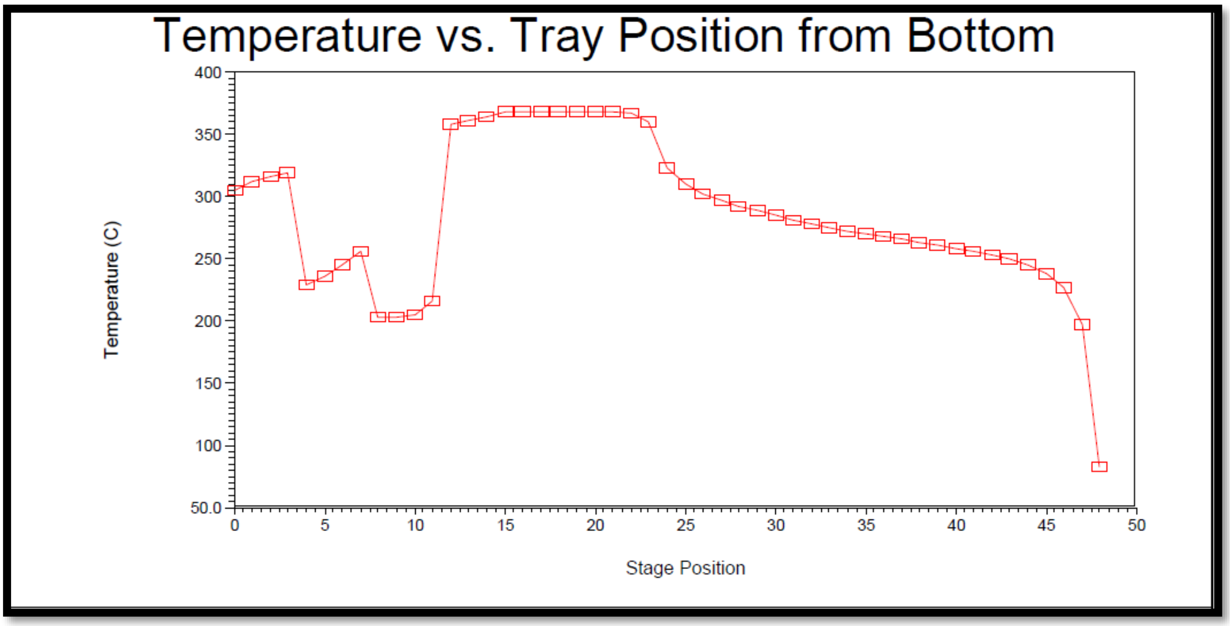


Figure 4.12: Temperature profile (Blend 2)

The quantity of liquid flow and Vapour flow shown in (figure 4.13)

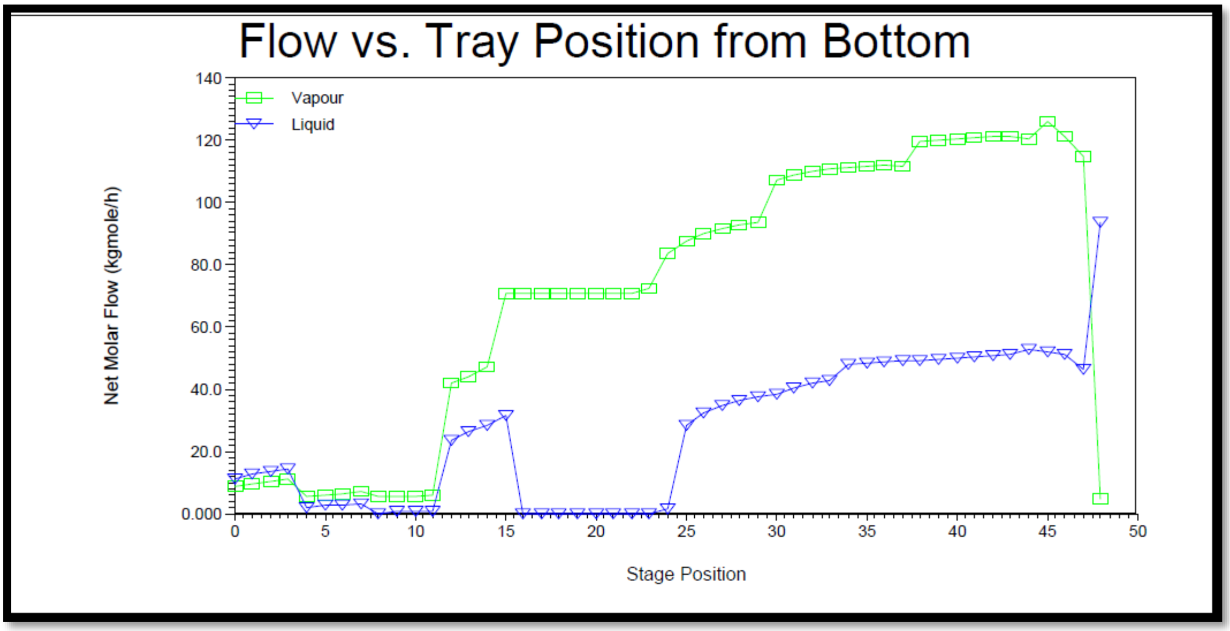


Figure 4.13: Net flow profile (Blend 2)

Column properties (Density and Molecular weight) shown in (figure 4.14)

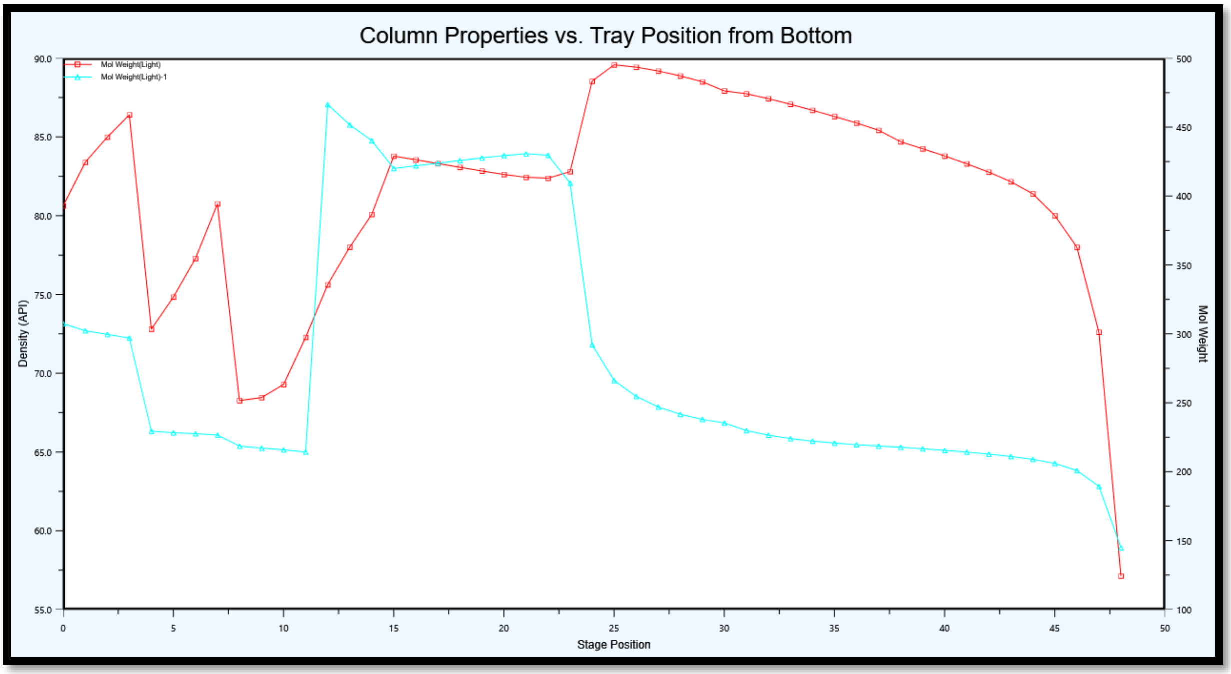


Figure 4.14: Correlation between density and molecular weight (Blend2)

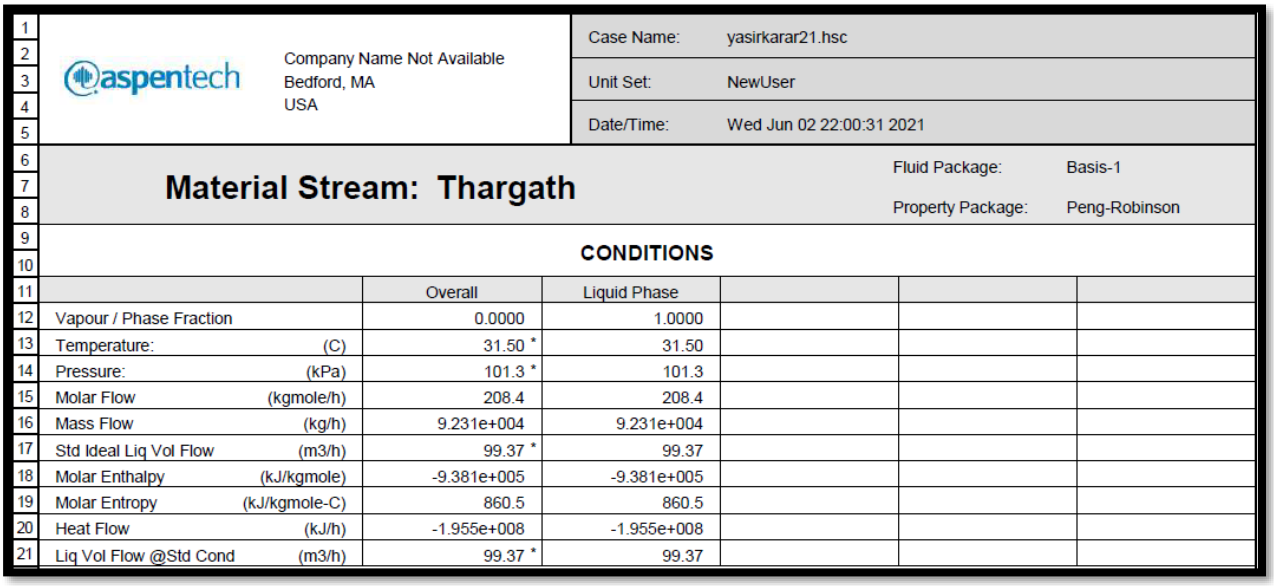
4.4.3 Case Study no.3 (Blend 3)

Pure Thargath crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.4: The Oil Distributions -Blend 3

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Light Naphtha	92.99	175	0.017
Heavy Naphtha	175	180	0.001
Kerosene	180	260	0.049
AGO	260	380	0.175
Residue	380	1027	0.758

(Figure 4.15) showing the Conditions of pure Thargath crude before entering the main distillation column.



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Company Name Not Available
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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:00:31 2021

Material Stream: Thargath

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

	Overall	Liquid Phase			
Vapour / Phase Fraction	0.0000	1.0000			
Temperature: (C)	31.50 *	31.50			
Pressure: (kPa)	101.3 *	101.3			
Molar Flow (kgmole/h)	208.4	208.4			
Mass Flow (kg/h)	9.231e+004	9.231e+004			
Std Ideal Liq Vol Flow (m3/h)	99.37 *	99.37			
Molar Enthalpy (kJ/kgmole)	-9.381e+005	-9.381e+005			
Molar Entropy (kJ/kgmole-C)	860.5	860.5			
Heat Flow (kJ/h)	-1.955e+008	-1.955e+008			
Liq Vol Flow @Std Cond (m3/h)	99.37 *	99.37			

Figure 4.15: Conditions of Thargath stream (Blend 3)

The following figures showing TBP curve (Figure 4.16) and Cuts Distribution (Figure 4.17).

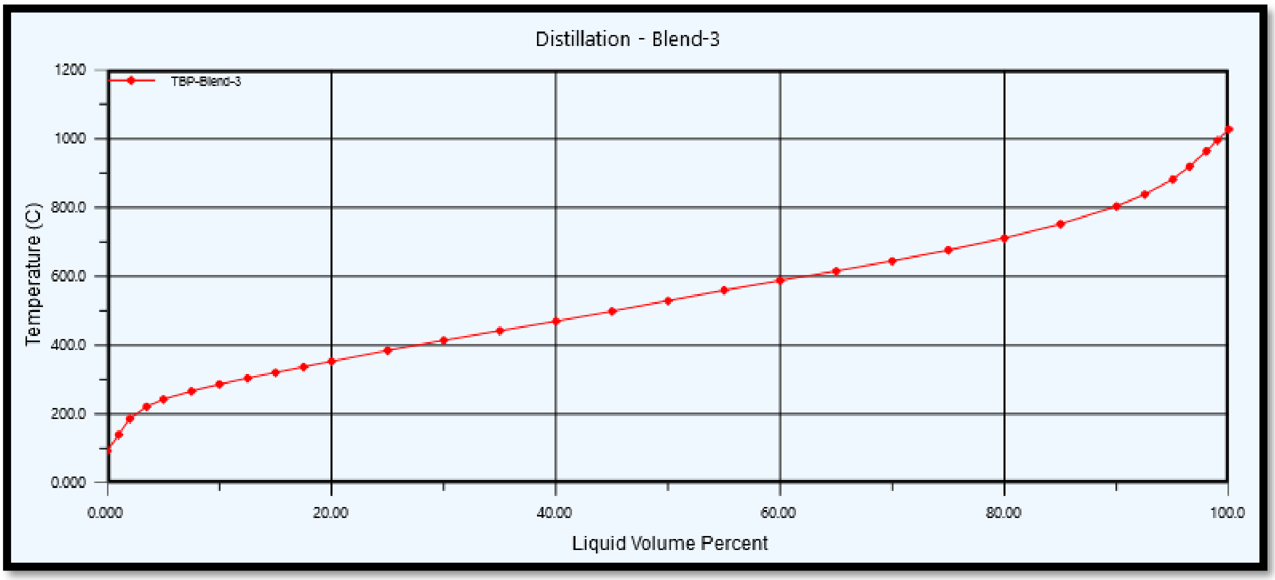


Figure 4.16: True Boiling Point Distillation Curve-Blend 3

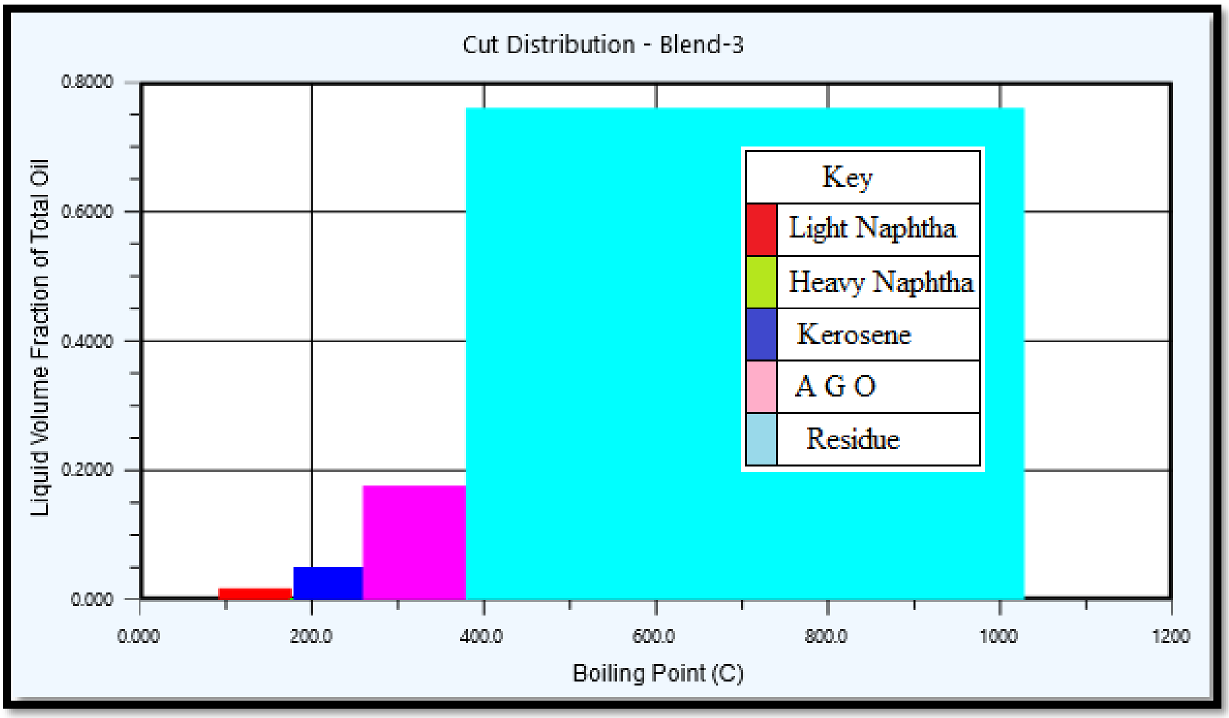


Figure 4.17: Cut Distribution-Blend 3

(Figure 4.18) showing the increasing of pressure during process

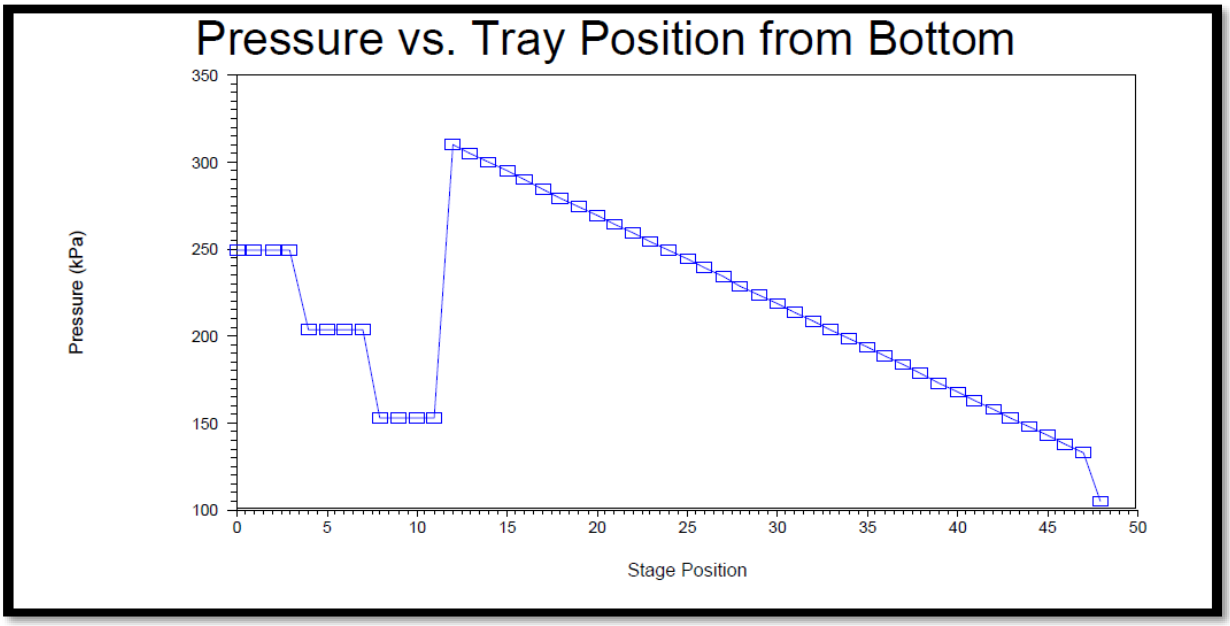


Figure 4.18: Pressure profile (Blend 3)

The change of temperature from tray to another shown in (figure 4.19)

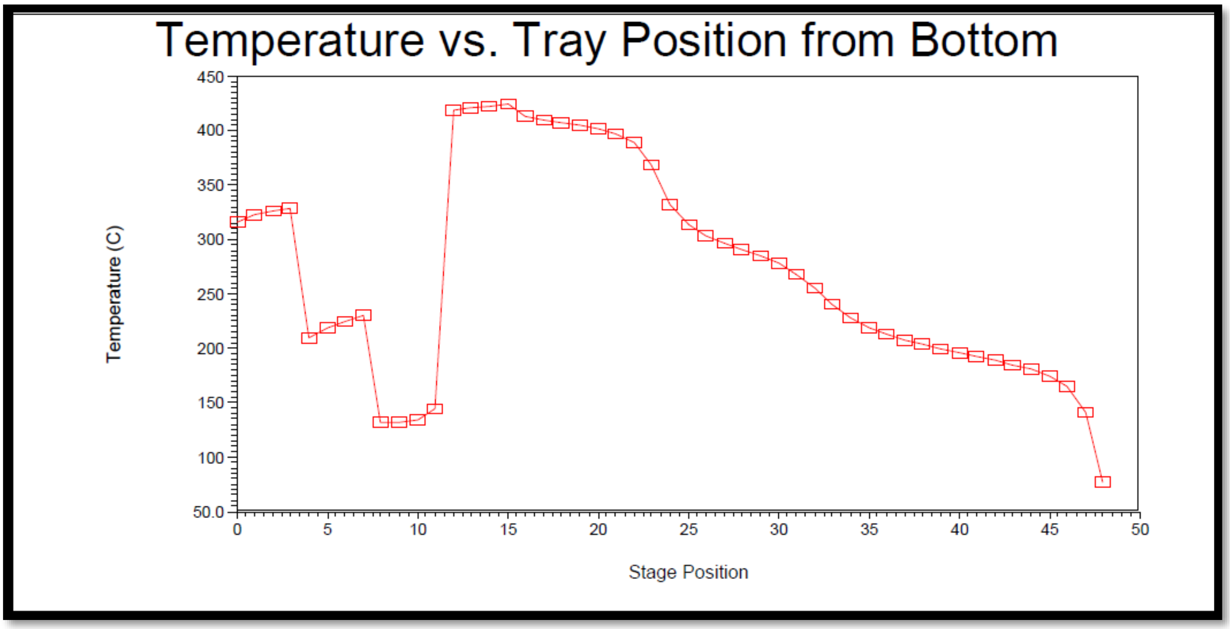


Figure 4.19: Temperature profile (Blend 3)

The quantity of liquid flow and Vapour flow shown in (figure 4.20)

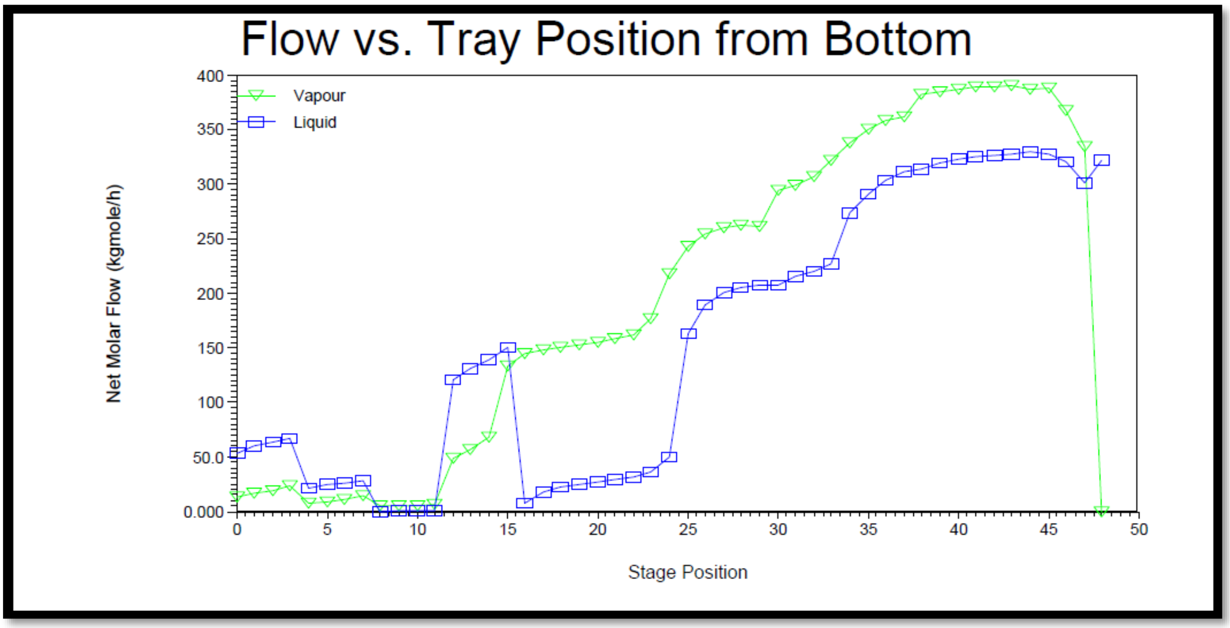


Figure 4.20: Net flow profile (Blend 3)

Column properties (Density and Molecular weight) shown in (figure 4.21)

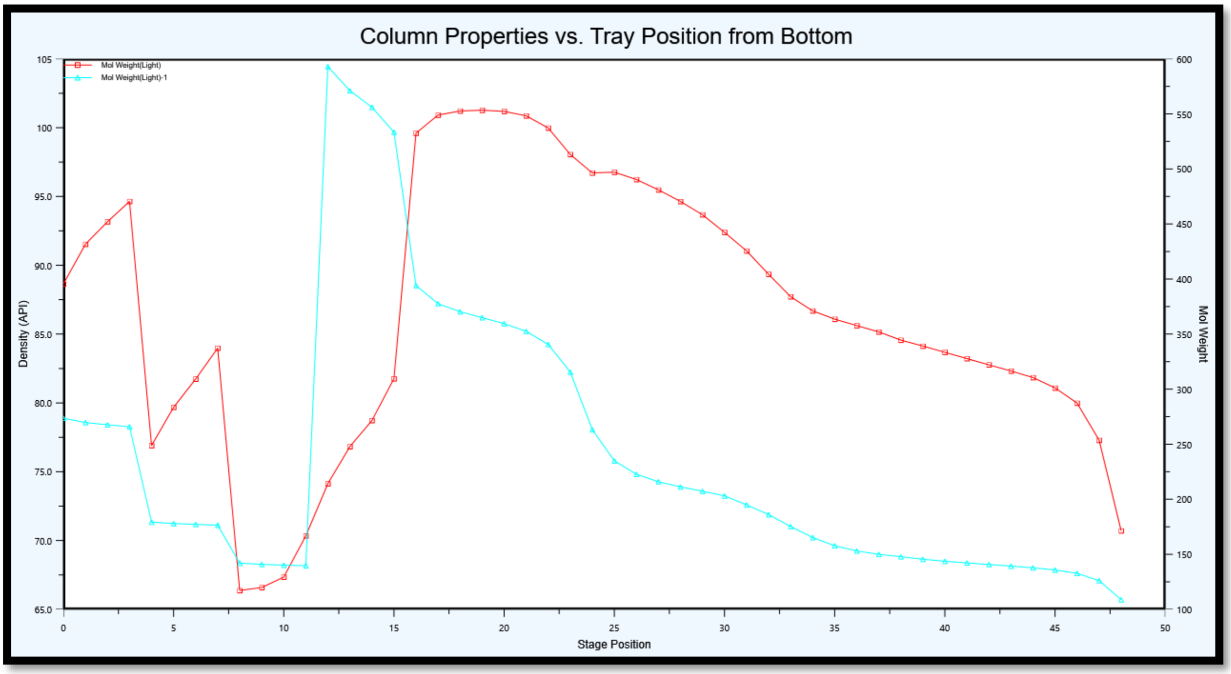


Figure 4.21: Correlation between density and molecular weight (Blend 3)

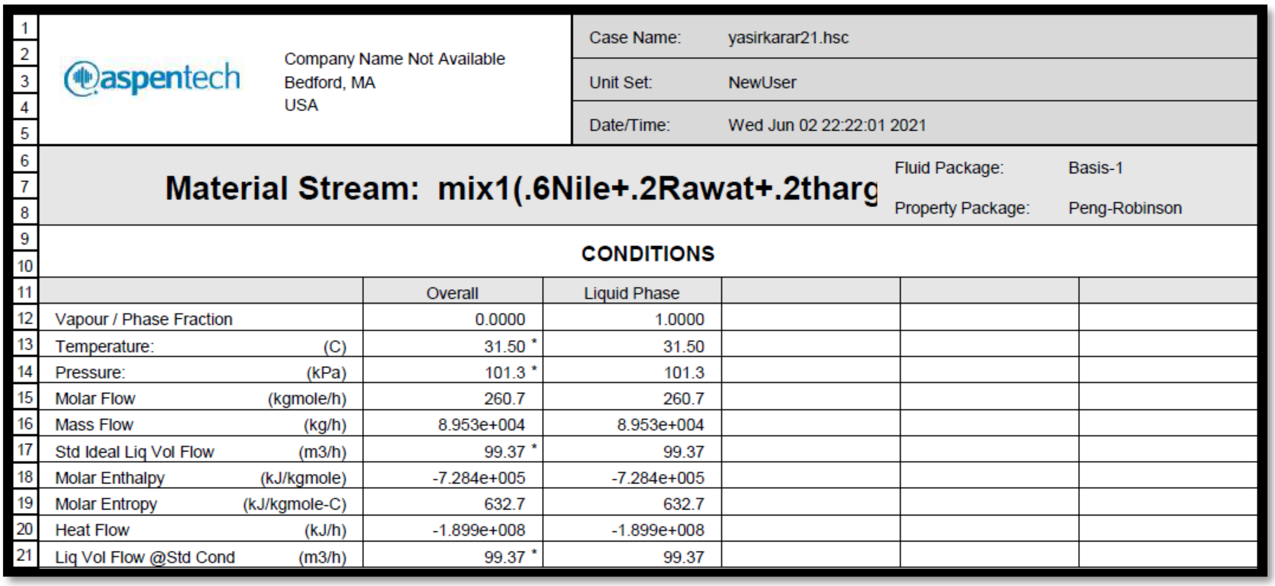
4.4.4 Case Study no.4 (Blend 4)

(60%Nile Blend+20%Rawat+20%Thargath) as defined Mix1 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.5: The Oil Distributions -Blend 4

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	36.98	59	0.013
Light Naphtha	59	175	0.061
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.075
AGO	260	380	0.191
Residue	380	1005	0.657

(Figure 4.22) showing the Conditions of Mix 1(blend 4) before entering the main distillation column.



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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:22:01 2021

Material Stream: mix1(.6Nile+.2Rawat+.2tharg

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

		Overall	Liquid Phase			
Vapour / Phase Fraction		0.0000	1.0000			
Temperature:	(C)	31.50 *	31.50			
Pressure:	(kPa)	101.3 *	101.3			
Molar Flow	(kgmole/h)	260.7	260.7			
Mass Flow	(kg/h)	8.953e+004	8.953e+004			
Std Ideal Liq Vol Flow	(m3/h)	99.37 *	99.37			
Molar Enthalpy	(kJ/kgmole)	-7.284e+005	-7.284e+005			
Molar Entropy	(kJ/kgmole-C)	632.7	632.7			
Heat Flow	(kJ/h)	-1.899e+008	-1.899e+008			
Liq Vol Flow @Std Cond	(m3/h)	99.37 *	99.37			

Figure 4.22: Conditions of Mix 1 stream (Blend 4)

The following figures showing TBP curve (Figure 4.23) and Cuts Distribution (Figure 4.24).

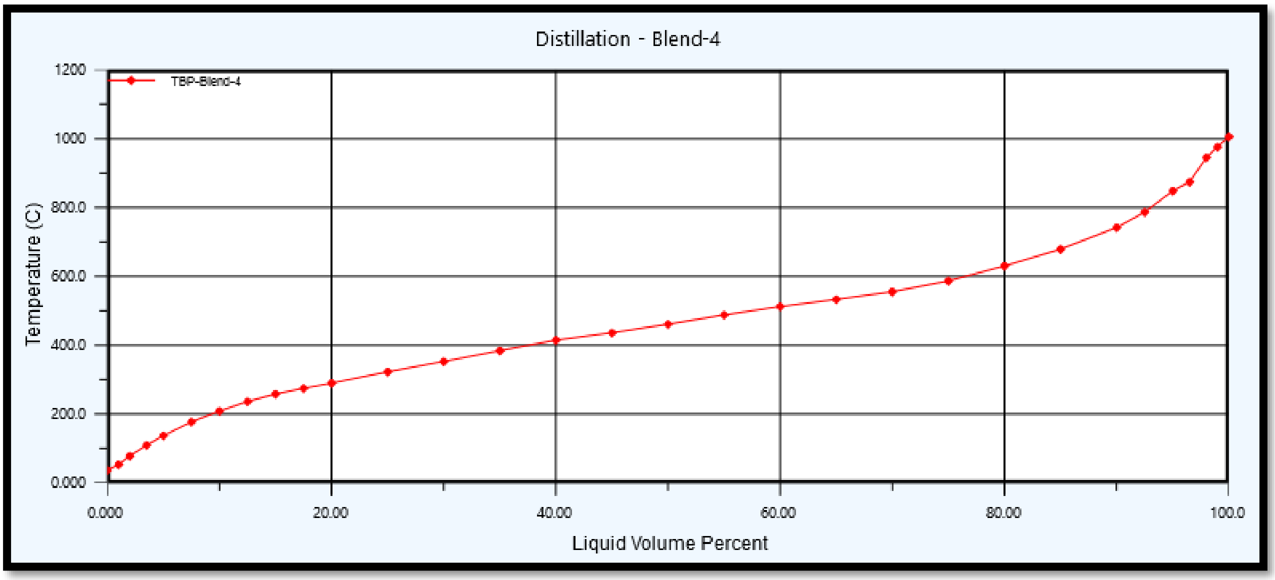


Figure 4.23: True Boiling Point Distillation Curve-Blend 4

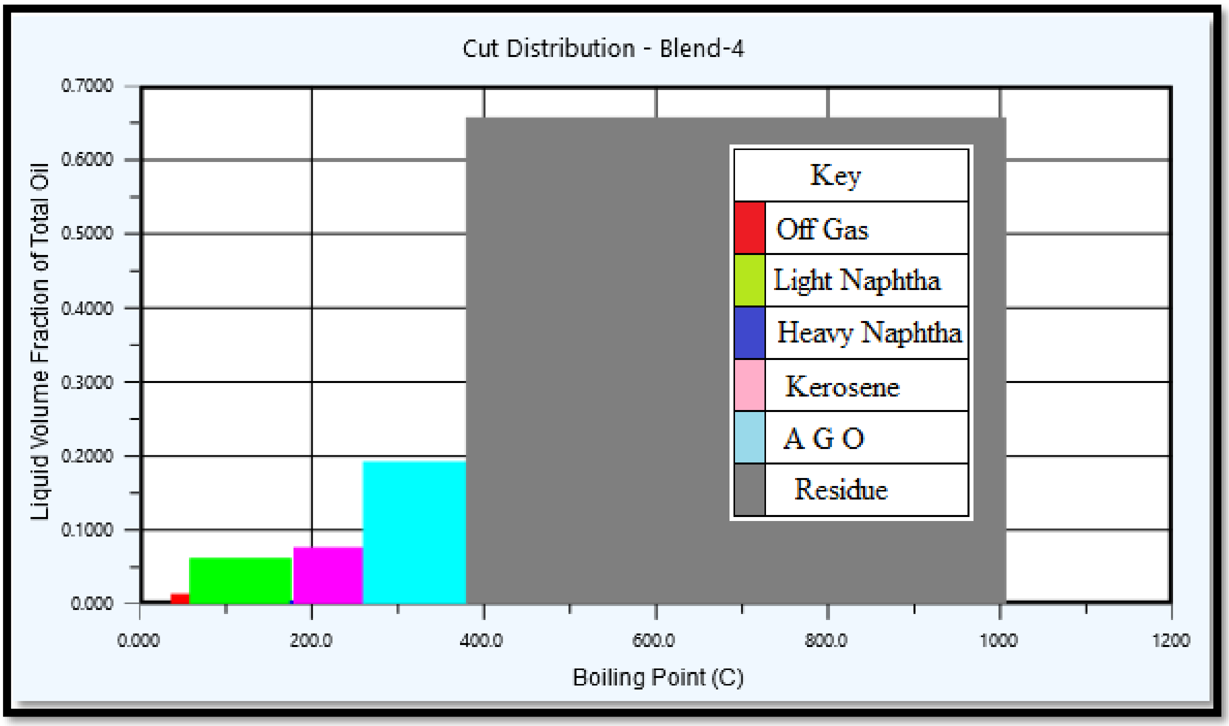


Figure 4.24: Cut Distribution-Blend 4

(Figure 4.25) showing the increasing of pressure during process

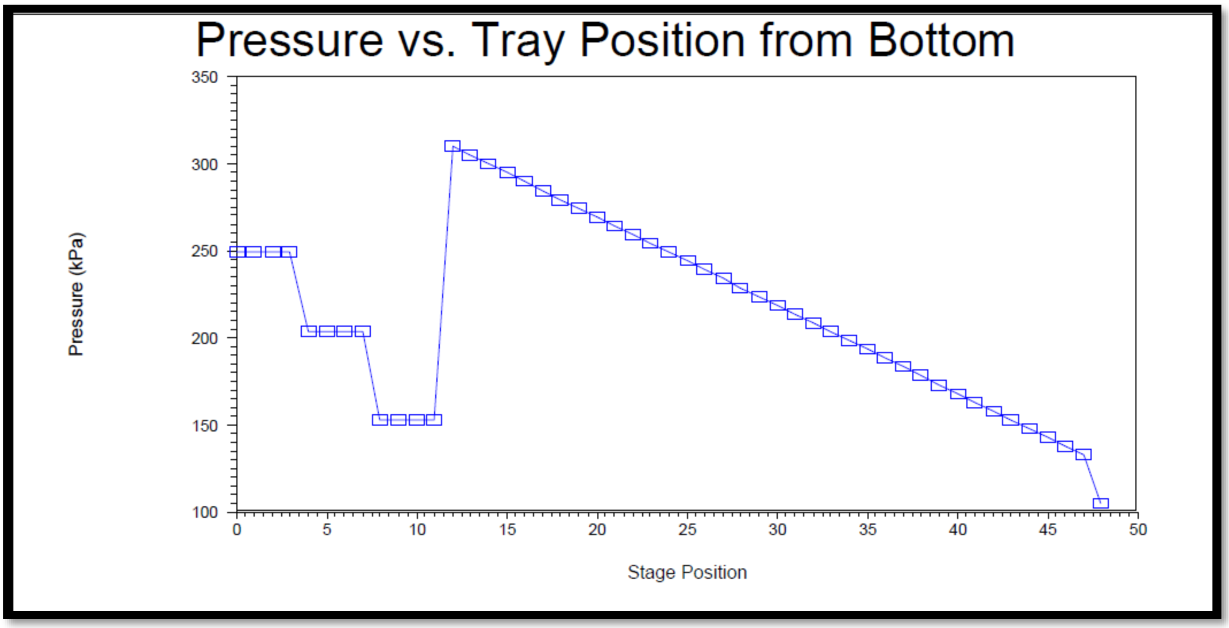


Figure 4.25: Pressure profile (Blend 4)

The change of temperature from tray to another shown in (figure 4.26)

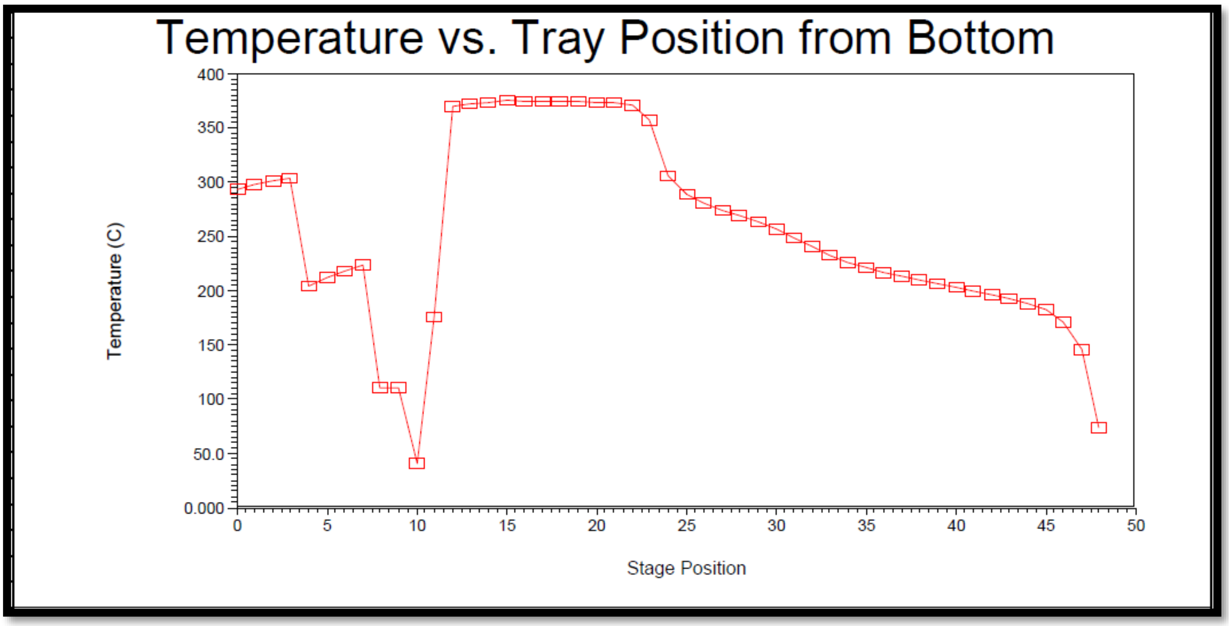


Figure 4.26: Temperature profile (Blend 4)

The quantity of liquid flow and Vapour flow shown in (figure 4.27)

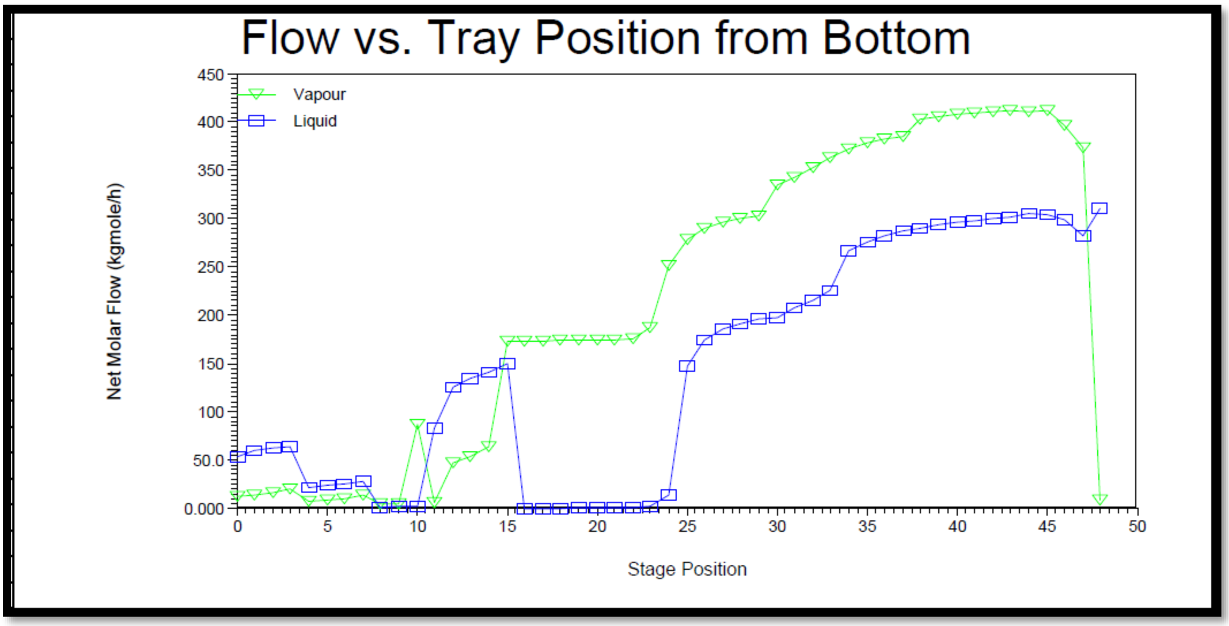


Figure 4.27: Net flow profile (Blend 4)

Column properties (Density and Molecular weight) shown in (figure 4.28)

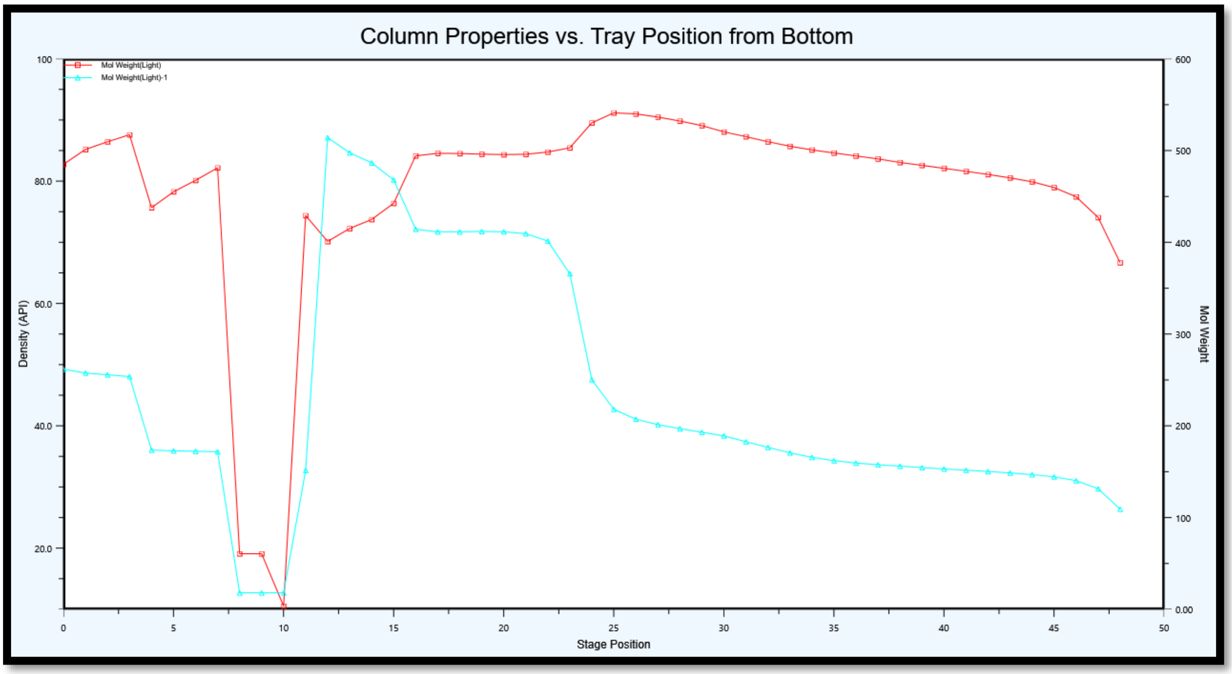


Figure 4.28: Correlation between density and molecular weight (Blend 4)

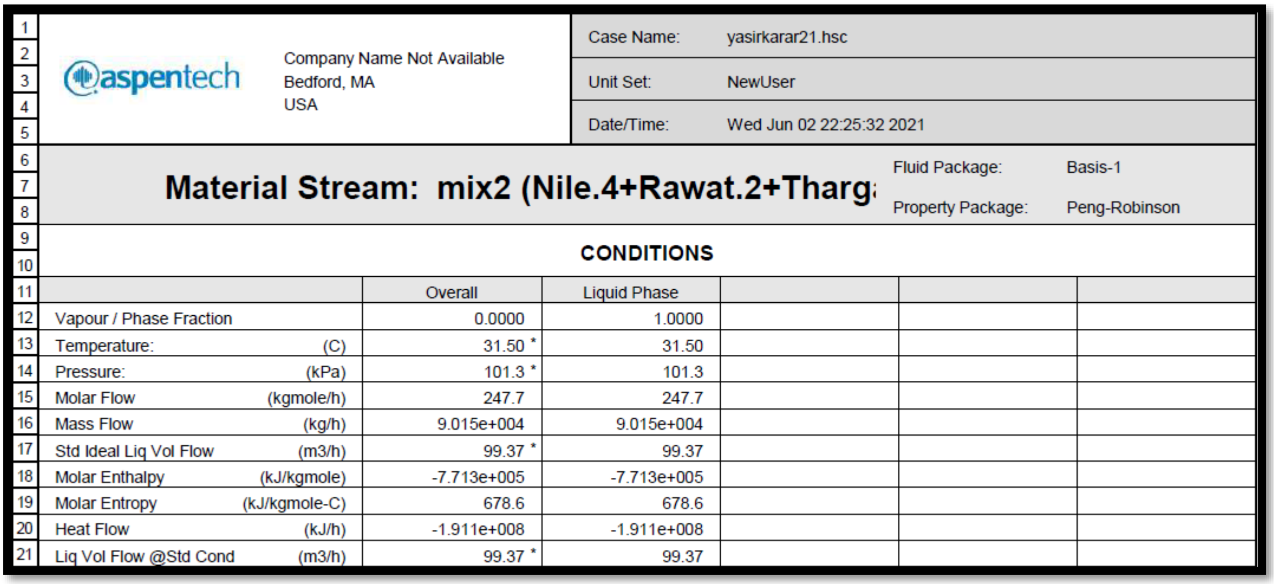
4.4.5 Case Study no.5 (Blend 5)

(40%Nile Blend+20%Rawat+40%Thargath) as defined Mix2 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.6: The Oil Distributions -Blend 5

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	41.34	59	0.010
Light Naphtha	59	175	0.048
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.068
AGO	260	380	0.191
Residue	380	1007	0.680

(Figure 4.29) showing the Conditions of Mix 2(blend 5) before entering the main distillation column.



Company Name Not Available
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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:25:32 2021

Material Stream: mix2 (Nile.4+Rawat.2+Tharg

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

	Overall	Liquid Phase
Vapour / Phase Fraction	0.0000	1.0000
Temperature: (C)	31.50 *	31.50
Pressure: (kPa)	101.3 *	101.3
Molar Flow (kgmole/h)	247.7	247.7
Mass Flow (kg/h)	9.015e+004	9.015e+004
Std Ideal Liq Vol Flow (m3/h)	99.37 *	99.37
Molar Enthalpy (kJ/kgmole)	-7.713e+005	-7.713e+005
Molar Entropy (kJ/kgmole-C)	678.6	678.6
Heat Flow (kJ/h)	-1.911e+008	-1.911e+008
Liq Vol Flow @Std Cond (m3/h)	99.37 *	99.37

Figure 4.29: Conditions of Mix 2 stream (Blend 5)

The following figures showing TBP curve (Figure 4.30) and Cuts Distribution (Figure 4.31).

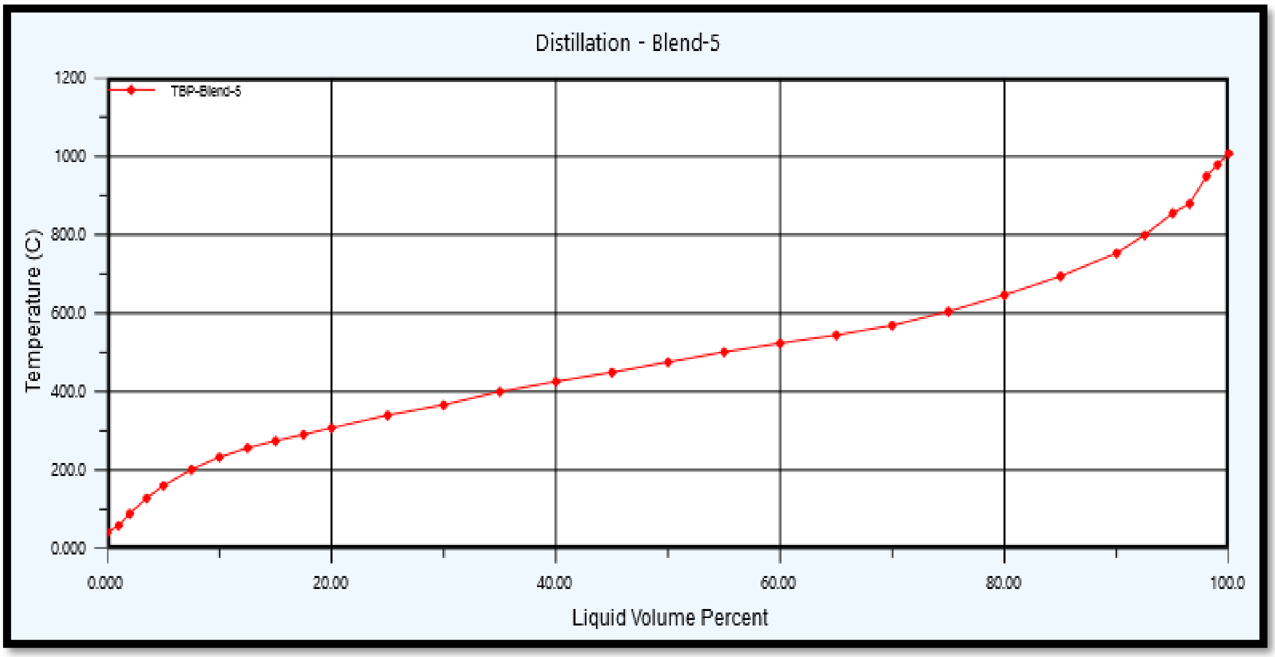


Figure 4.30: True Boiling Point Distillation Curve-Blend 5

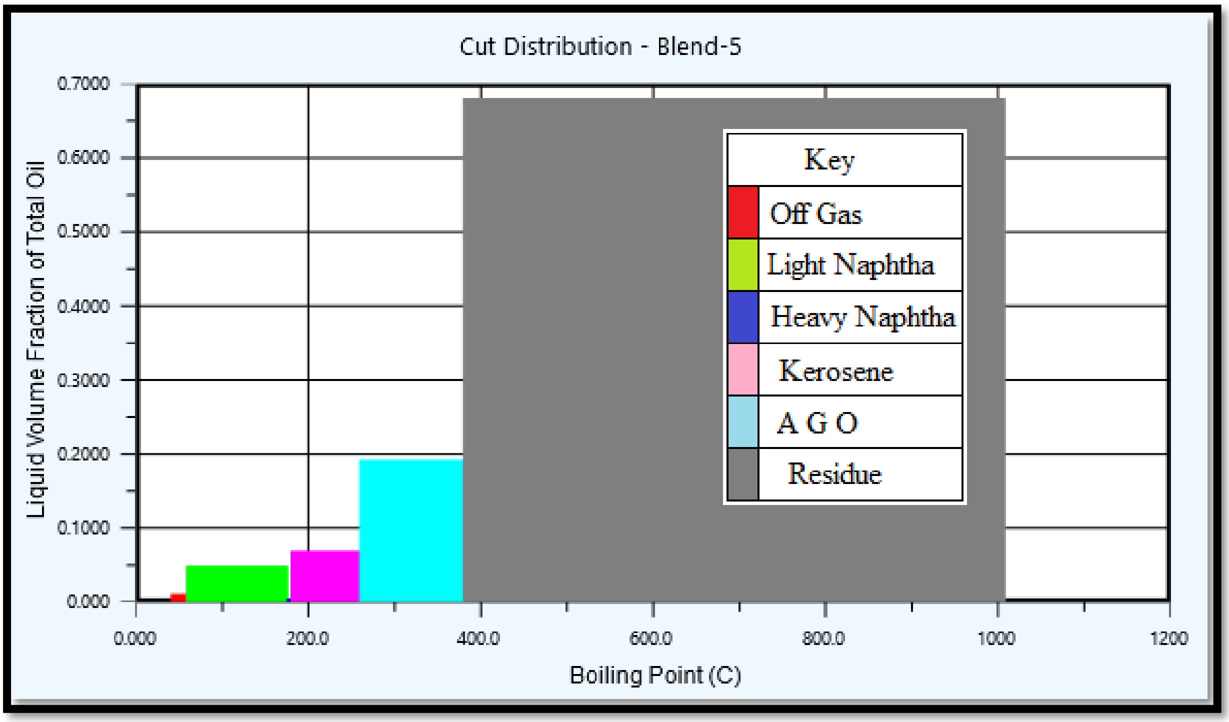


Figure 4.31: Cut Distribution-Blend 5

(Figure 4.32) showing the increasing of pressure during process

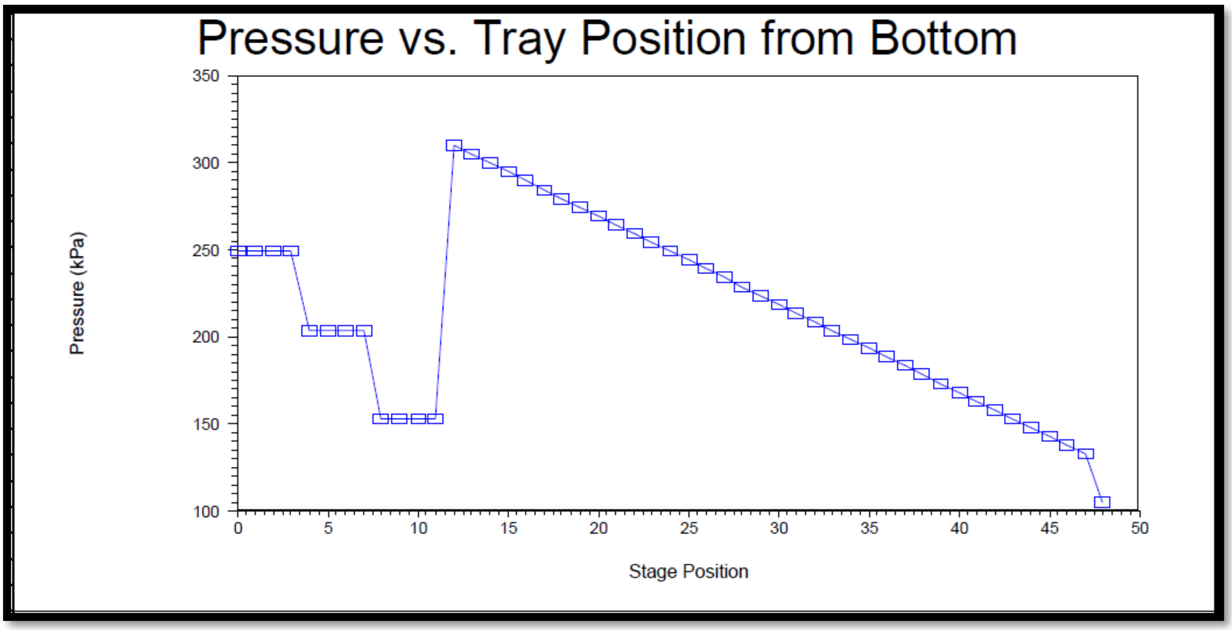


Figure 4.32: Pressure profile (Blend 5)

The change of temperature from tray to another shown in (figure 4.33)

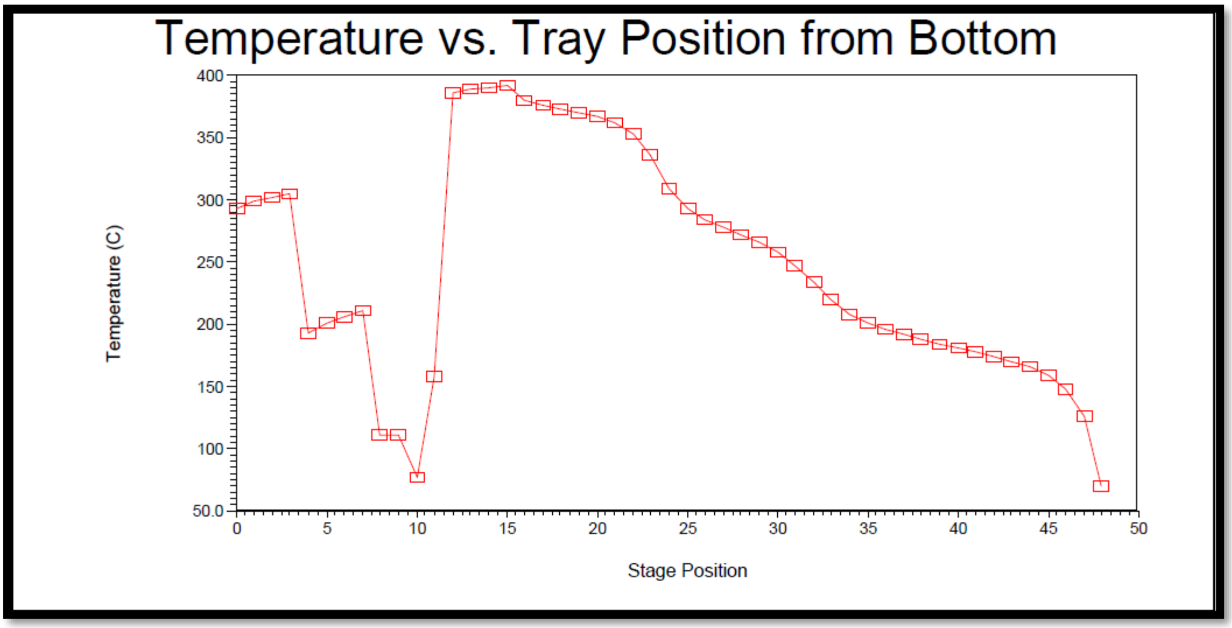


Figure 4.33: Temperature profile (Blend 5)

The quantity of liquid flow and Vapour flow shown in (figure 4.34)

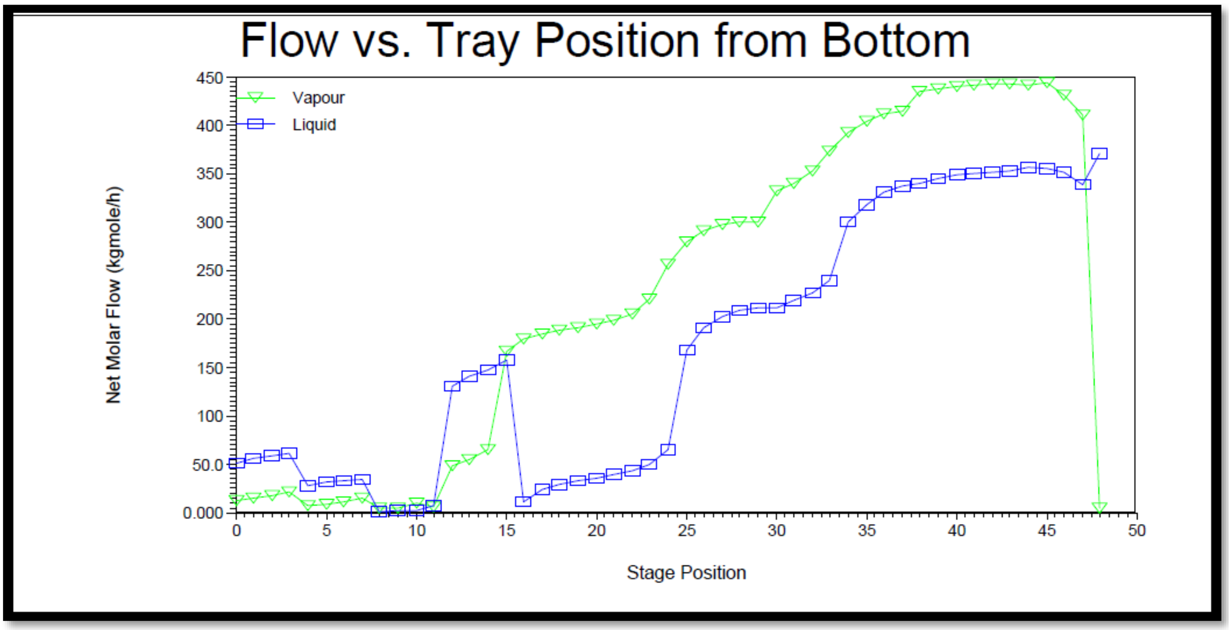


Figure 4.34: Net flow profile (Blend 5)

Column properties (Density and Molecular weight) shown in (figure 4.35)

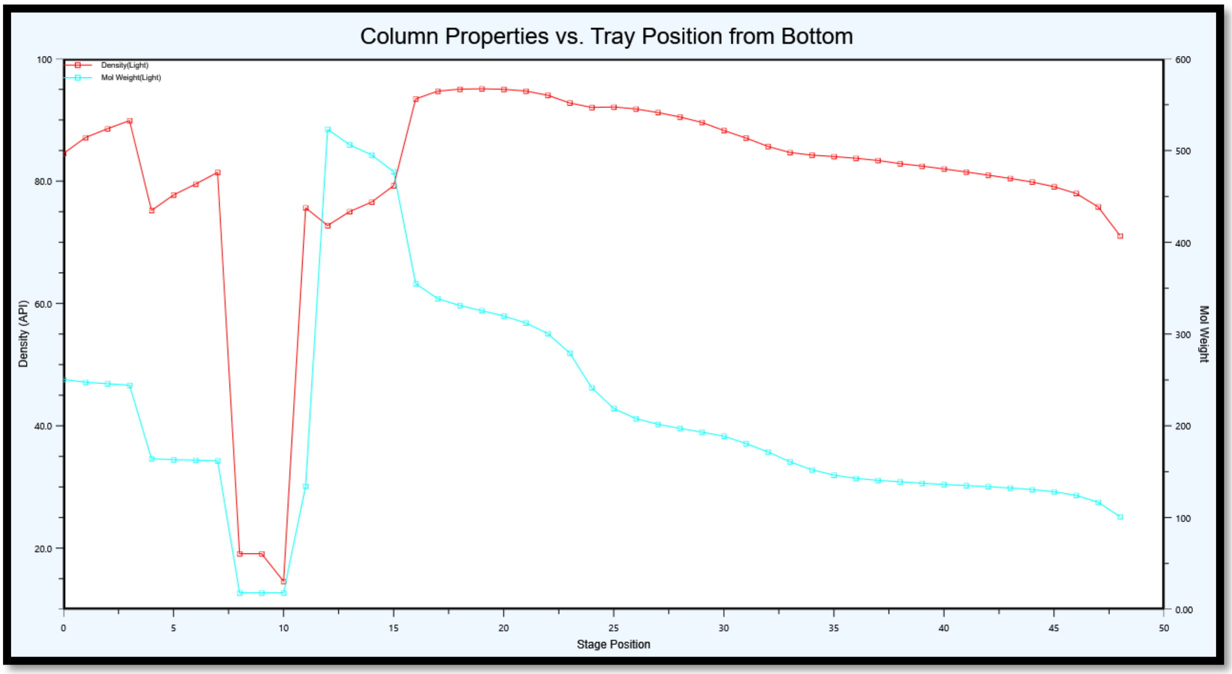


Figure 4.35: Correlation between density and molecular weight (Blend 5)

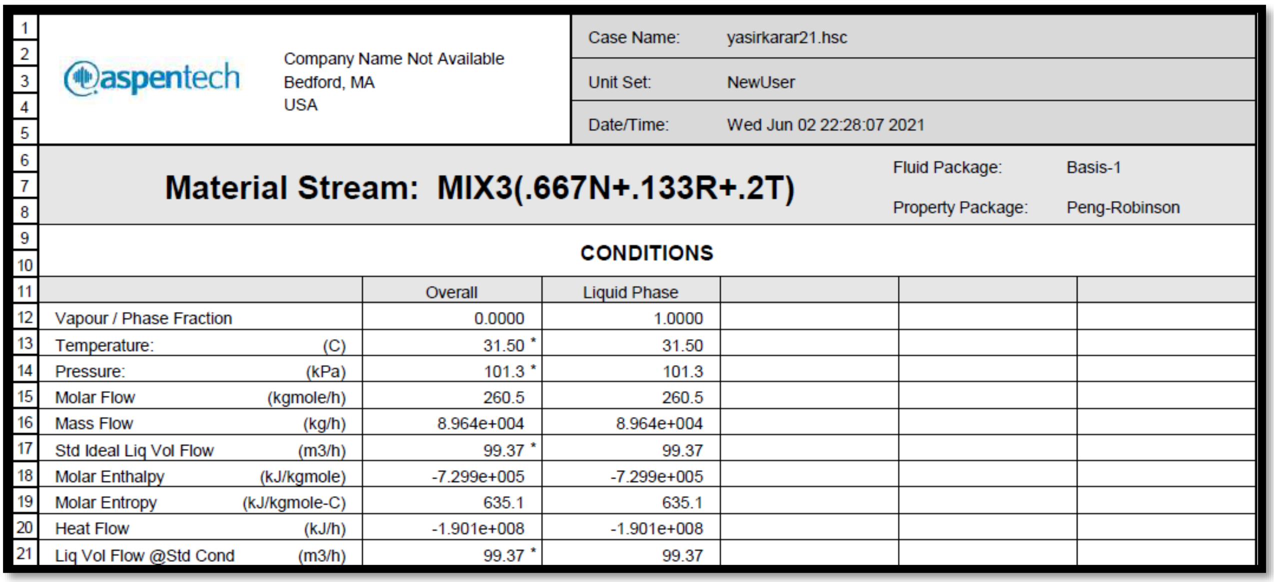
4.4.6 Case Study no.6 (Blend 6)

(66.7%Nile Blend+13.3%Rawat+20%Thargath) as defined Mix 3 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.7: The Oil Distributions -Blend 6

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	37.43	59	0.012
Light Naphtha	59	175	0.063
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.076
AGO	260	380	0.186
Residue	380	1009	0.659

(Figure 4.36) showing the Conditions of Mix 3(blend 6) before entering the main distillation column.



Company Name Not Available
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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:28:07 2021

Material Stream: MIX3(.667N+.133R+.2T)

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

	Overall	Liquid Phase			
Vapour / Phase Fraction	0.0000	1.0000			
Temperature: (C)	31.50 *	31.50			
Pressure: (kPa)	101.3 *	101.3			
Molar Flow (kgmole/h)	260.5	260.5			
Mass Flow (kg/h)	8.964e+004	8.964e+004			
Std Ideal Liq Vol Flow (m3/h)	99.37 *	99.37			
Molar Enthalpy (kJ/kgmole)	-7.299e+005	-7.299e+005			
Molar Entropy (kJ/kgmole-C)	635.1	635.1			
Heat Flow (kJ/h)	-1.901e+008	-1.901e+008			
Liq Vol Flow @Std Cond (m3/h)	99.37 *	99.37			

Figure 4.36: Conditions of Mix 3 stream (Blend 6)

The following figures showing TBP curve (Figure 4.37) and Cuts Distribution (Figure 4.38).

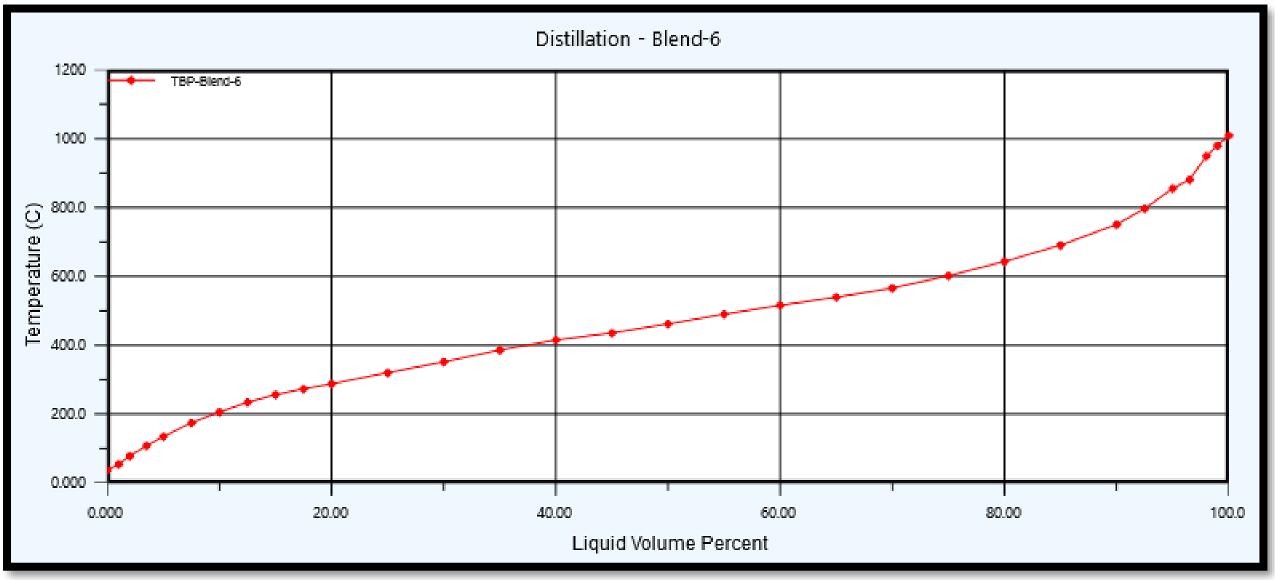


Figure 4.37: True Boiling Point Distillation Curve-Blend 6

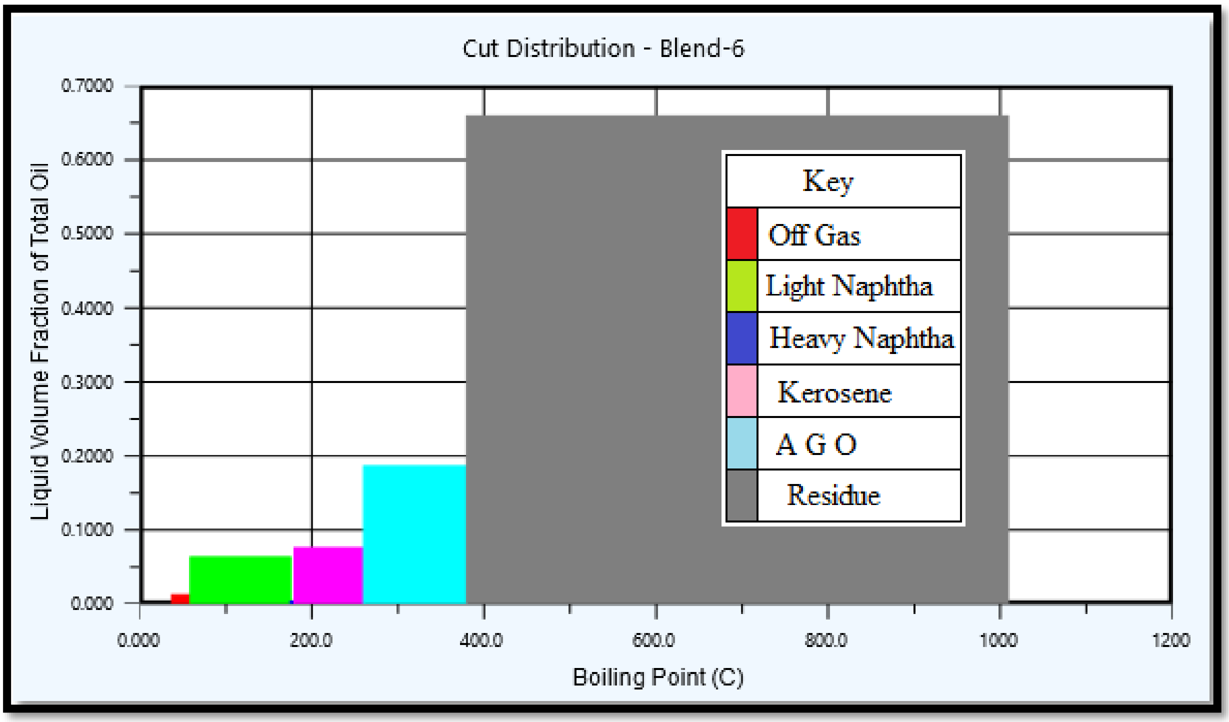


Figure 4.38: Cut Distribution-Blend 6

(Figure 4.39) showing the increasing of pressure during process

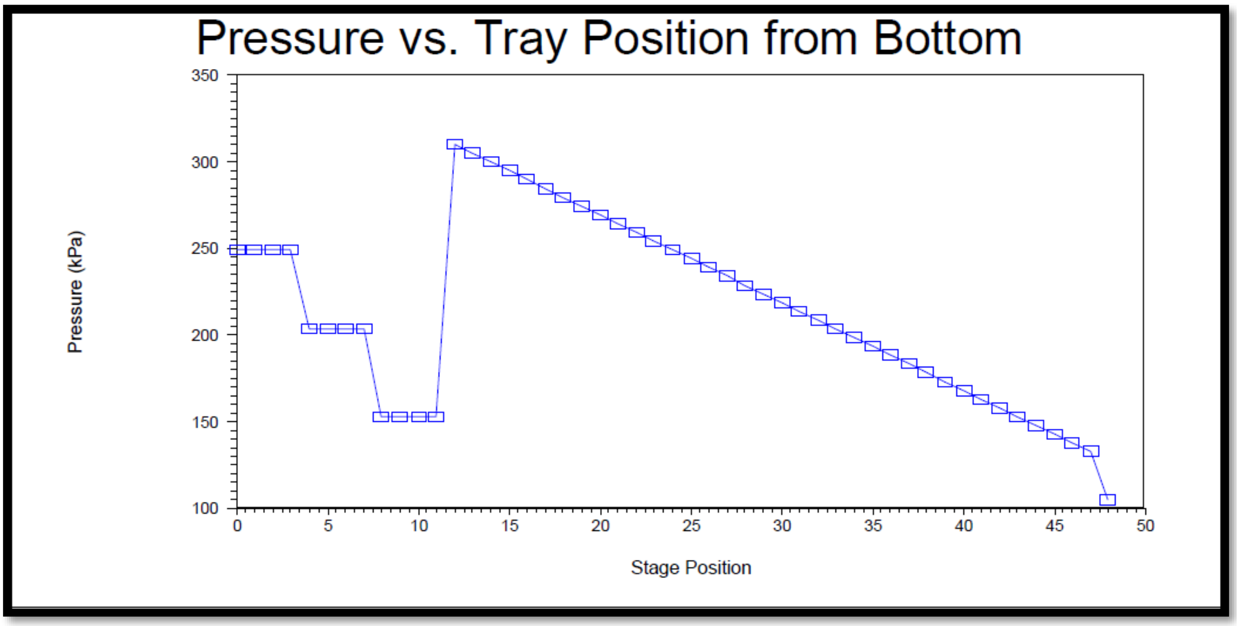


Figure 4.39: Pressure profile (Blend 6)

The change of temperature from tray to another shown in (figure 4.40)

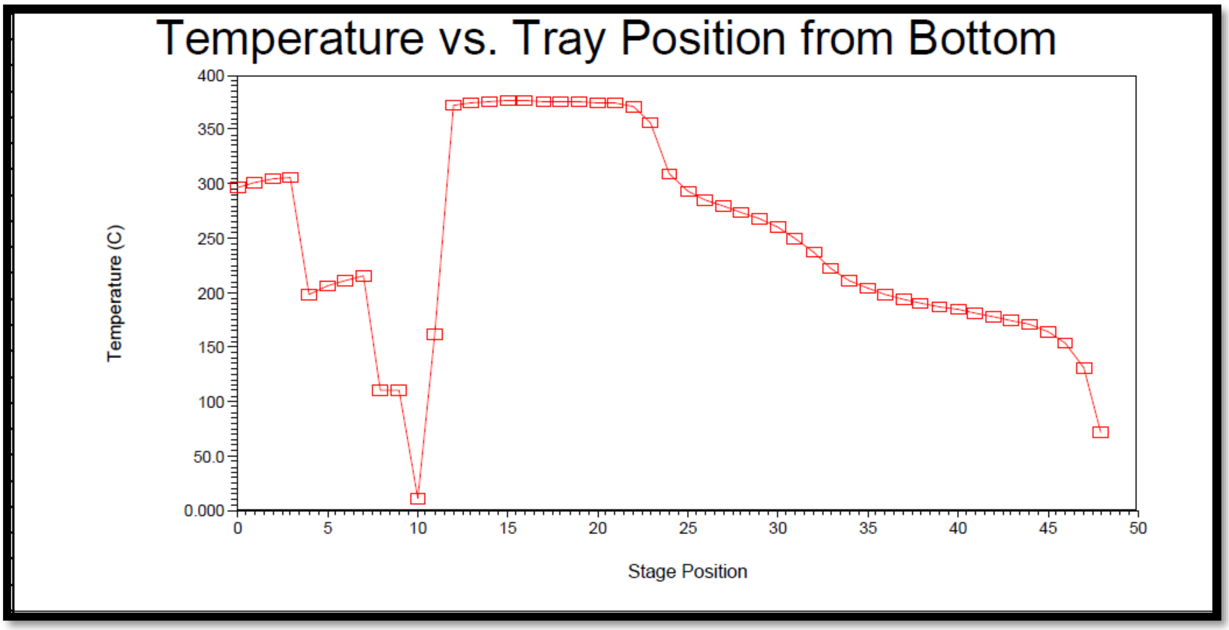


Figure 4.40: Temperature profile (Blend 6)

The quantity of liquid flow and Vapour flow shown in (figure 4.41)

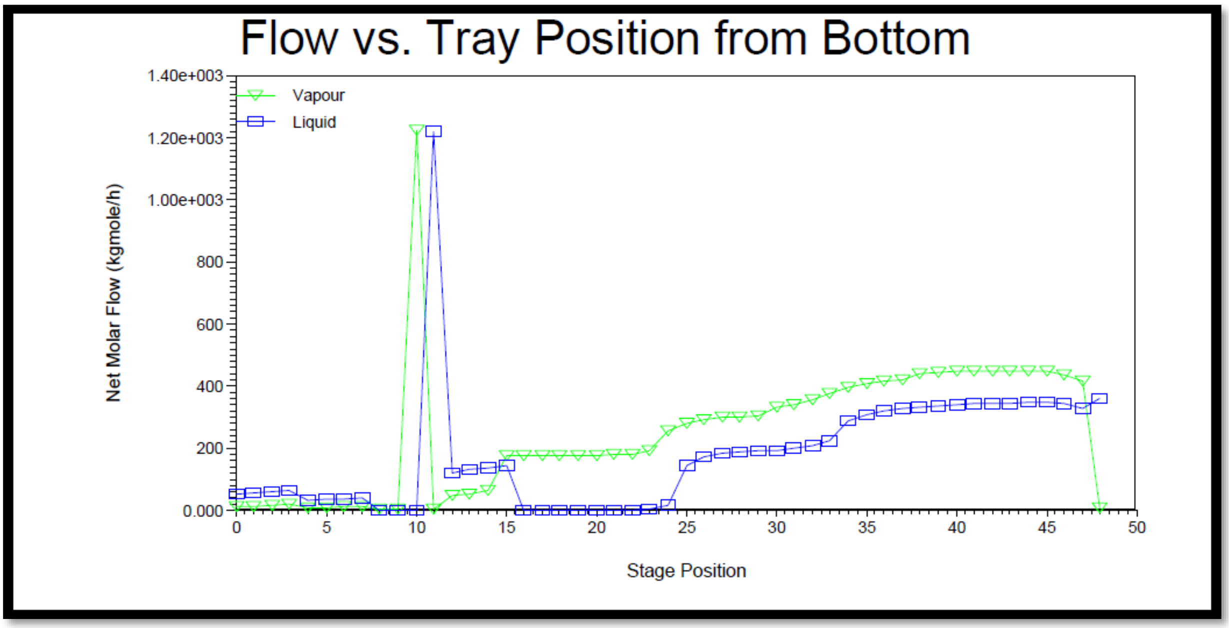


Figure 4.41: Net flow profile (Blend 6)

Column properties (Density and Molecular weight) shown in (figure 4.42)

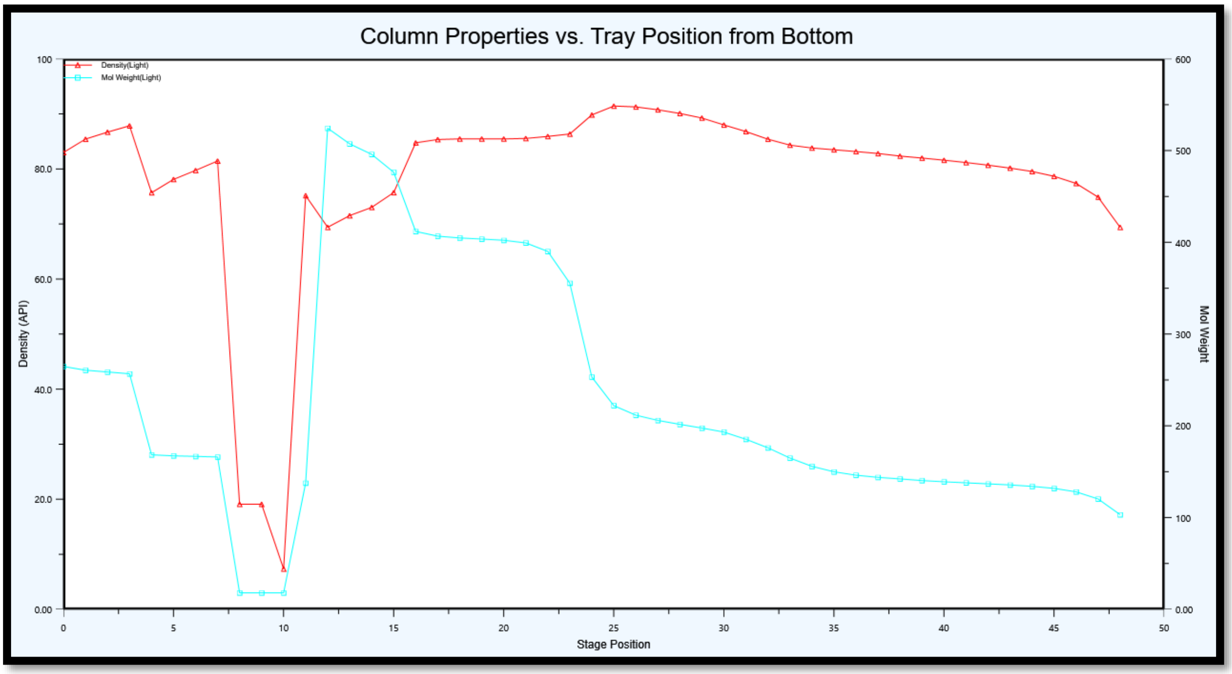


Figure 4.42: Correlation between density and molecular weight (Blend 6)

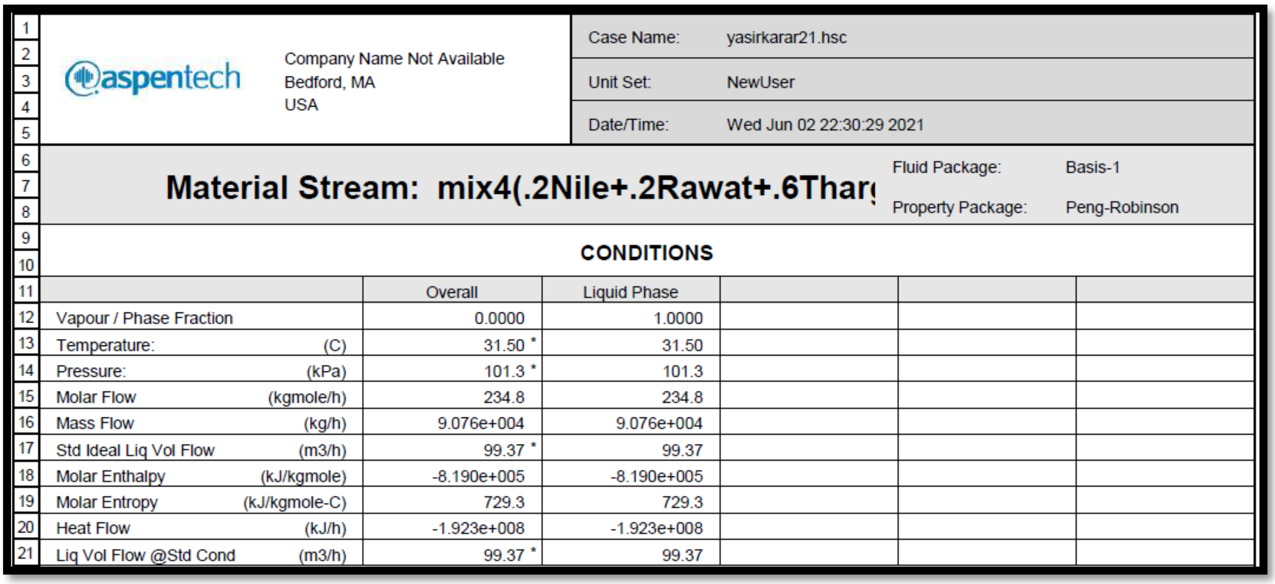
4.4.7 Case Study no.7 (Blend 7)

(20%Nile Blend+20%Rawat+60%Thargath) as defined Mix 4 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.8: The Oil Distributions -Blend 7

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	47.13	59	0.006
Light Naphtha	59	175	0.036
Heavy Naphtha	175	180	0.002
Kerosene	180	260	0.061
AGO	260	380	0.191
Residue	380	1010	0.703

(Figure 4.43) showing the Conditions of Mix 4(blend 7) before entering the main distillation column.



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Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:30:29 2021

Material Stream: mix4(.2Nile+.2Rawat+.6Thar

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

	Overall	Liquid Phase			
Vapour / Phase Fraction	0.0000	1.0000			
Temperature: (C)	31.50 *	31.50			
Pressure: (kPa)	101.3 *	101.3			
Molar Flow (kgmole/h)	234.8	234.8			
Mass Flow (kg/h)	9.076e+004	9.076e+004			
Std Ideal Liq Vol Flow (m3/h)	99.37 *	99.37			
Molar Enthalpy (kJ/kgmole)	-8.190e+005	-8.190e+005			
Molar Entropy (kJ/kgmole-C)	729.3	729.3			
Heat Flow (kJ/h)	-1.923e+008	-1.923e+008			
Liq Vol Flow @Std Cond (m3/h)	99.37 *	99.37			

Figure 4.43: Conditions of Mix 4 stream (Blend 7)

The following figures showing TBP curve (Figure 4.44) and Cuts Distribution (Figure 4.45).

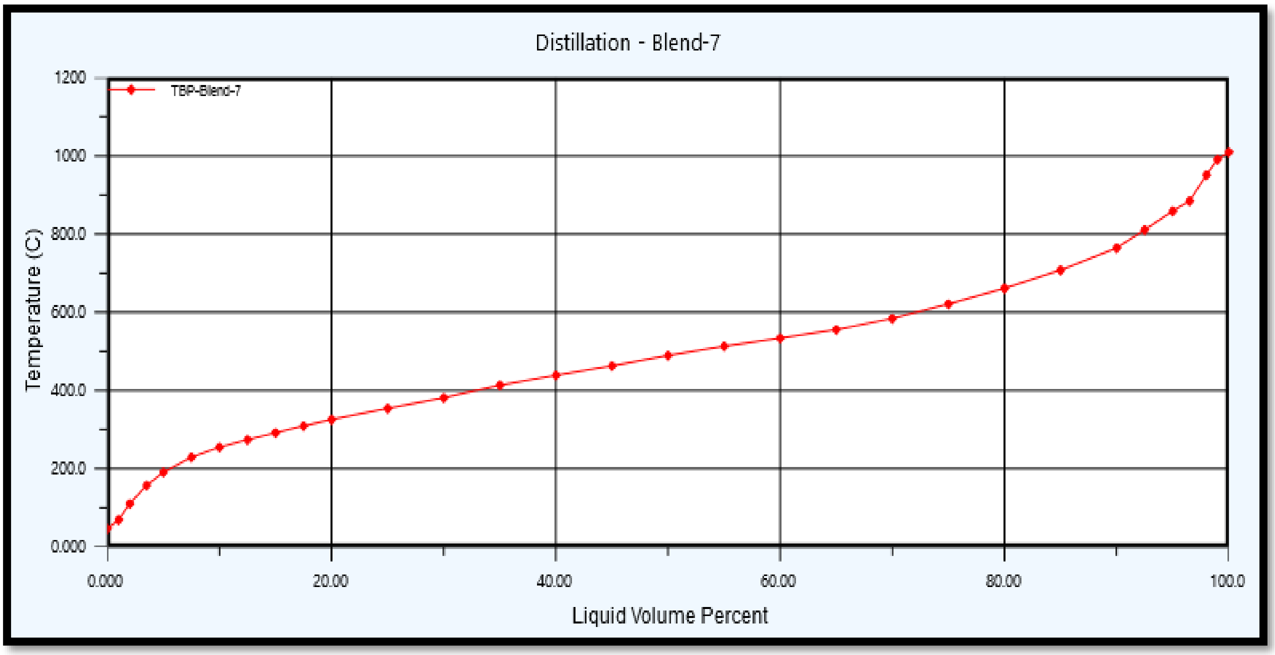


Figure 4.44: True Boiling Point Distillation Curve-Blend 7

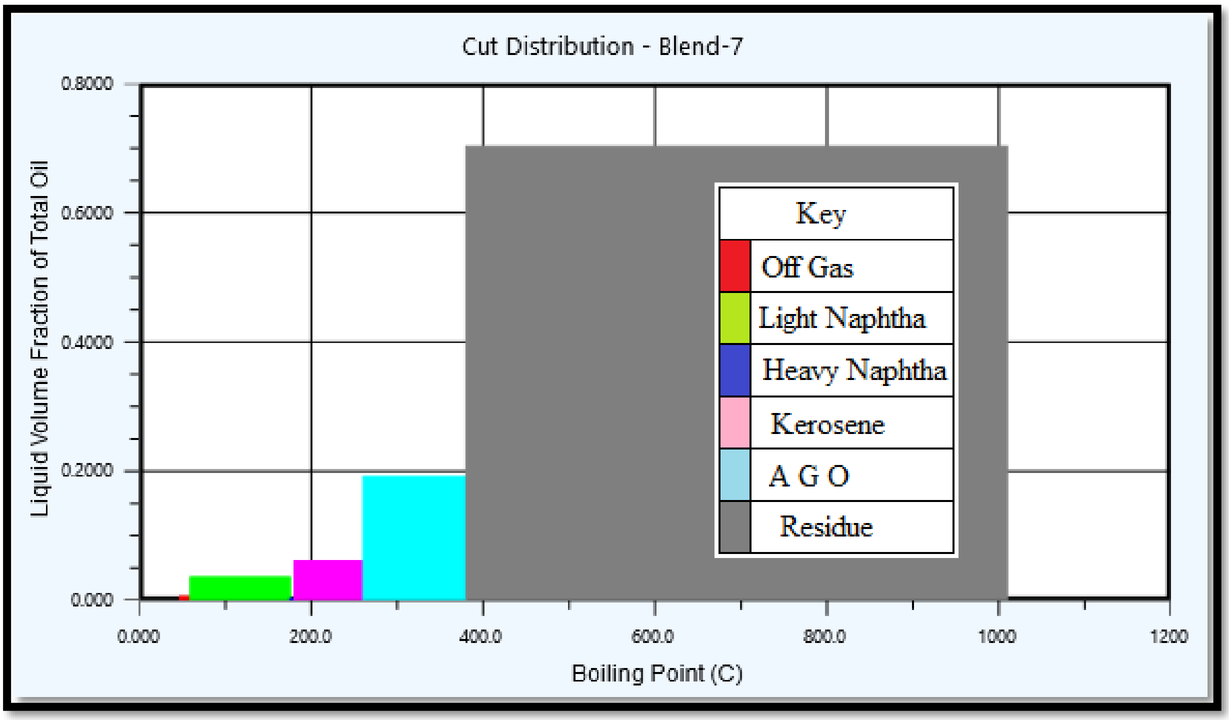


Figure 4.45: Cut Distribution-Blend 7

(Figure 4.46) showing the increasing of pressure during process

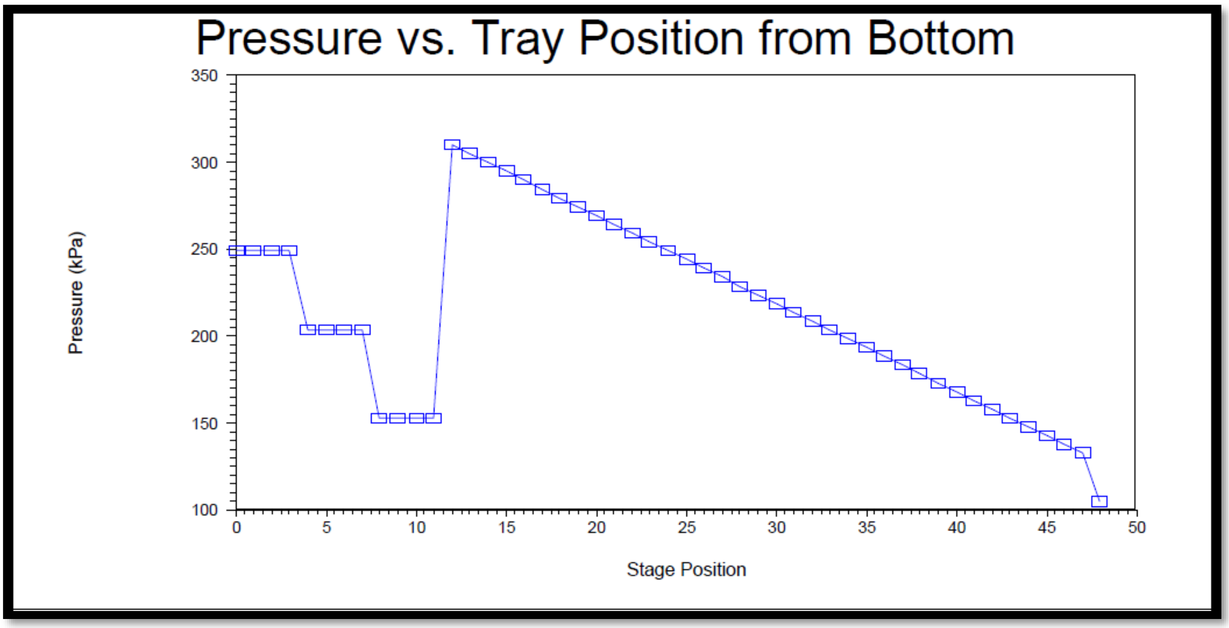


Figure 4.46: Pressure profile (Blend 7)

The change of temperature from tray to another shown in (figure 4.47)

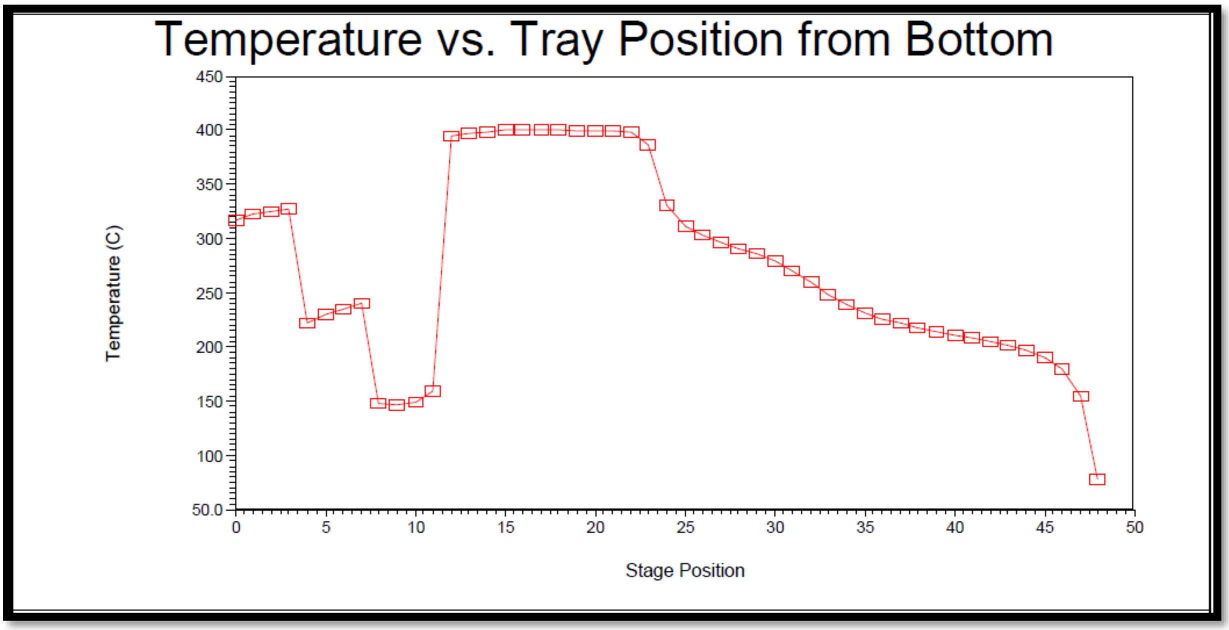


Figure 4.47: Temperature profile (Blend 7)

The quantity of liquid flow and Vapour flow shown in (figure 4.48)

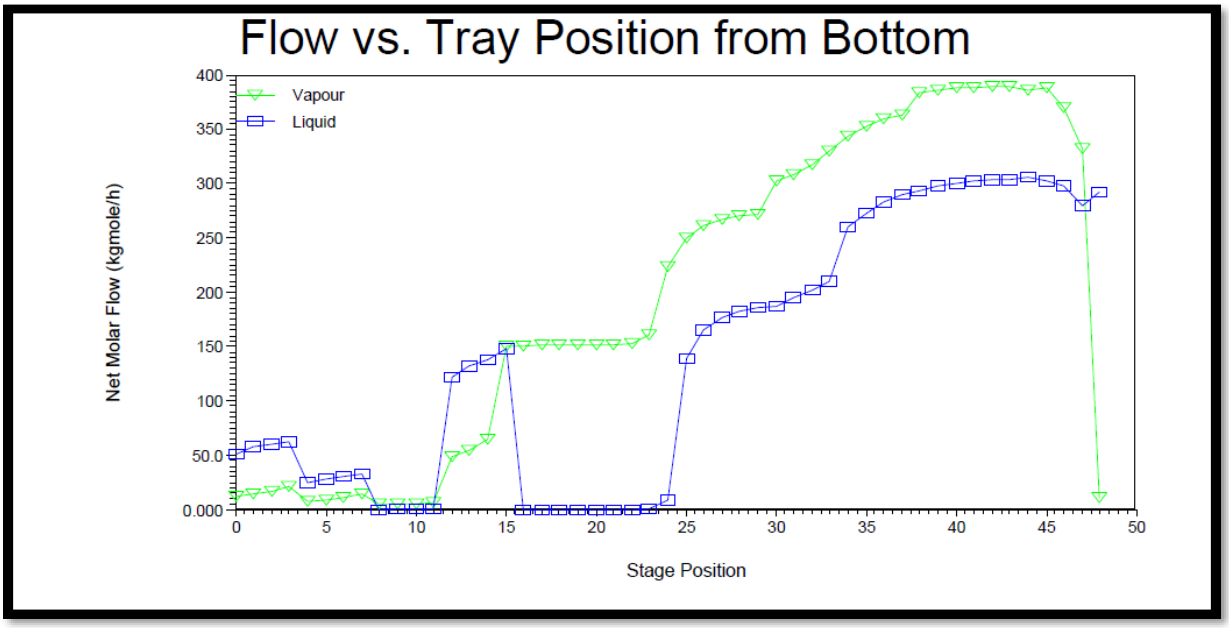


Figure 4.48: Net flow profile (Blend 7)

Column properties (Density and Molecular weight) shown in (figure 4.49)

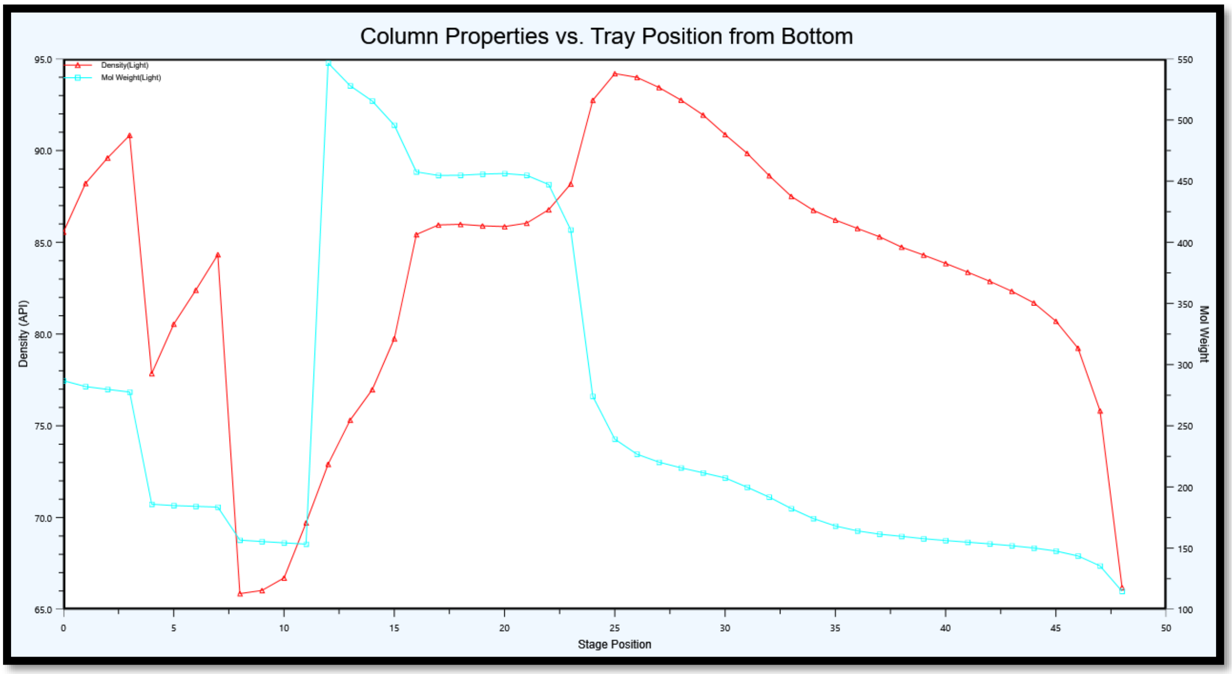


Figure 4.49: Correlation between density and molecular weight (Blend 7)

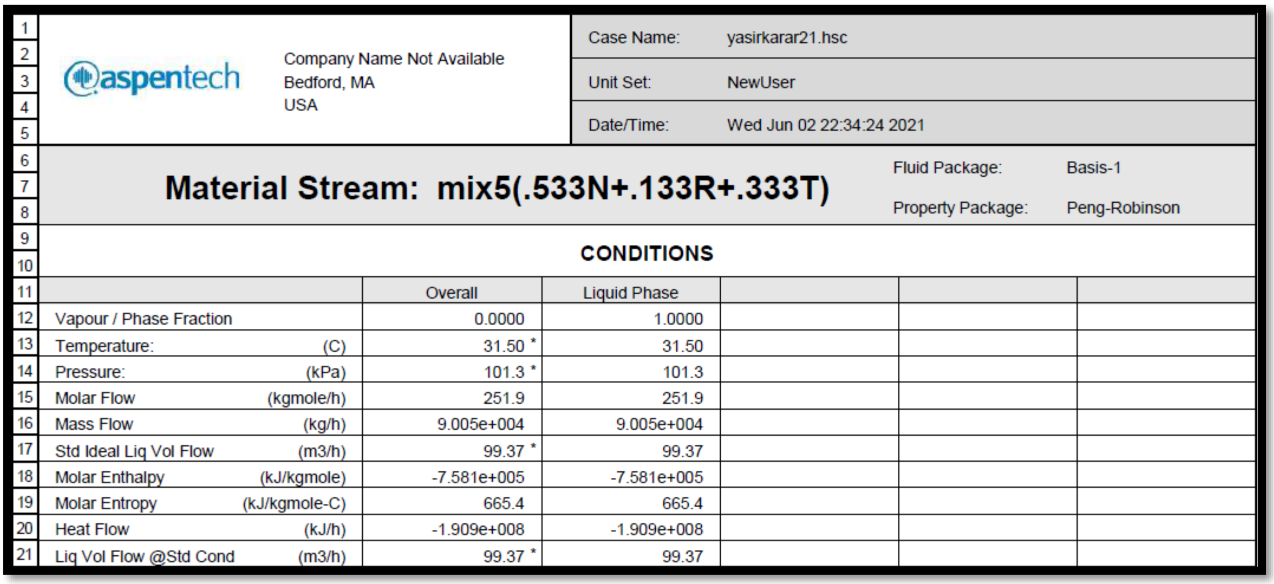
4.4.8 Case Study no.8 (Blend 8)

(53.33%Nile Blend+13.33%Rawat+33.33%Thargath) as defined Mix 5 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.9: The Oil Distributions -Blend 8

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	40.35	59	0.011
Light Naphtha	59	175	0.055
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.071
AGO	260	380	0.186
Residue	380	1010	0.675

(Figure 4.50) showing the Conditions of Mix 5(blend 8) before entering the main distillation column.



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Company Name Not Available
Bedford, MA
USA

Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:34:24 2021

Material Stream: mix5(.533N+.133R+.333T)

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

		Overall	Liquid Phase
Vapour / Phase Fraction		0.0000	1.0000
Temperature:	(C)	31.50 *	31.50
Pressure:	(kPa)	101.3 *	101.3
Molar Flow	(kgmole/h)	251.9	251.9
Mass Flow	(kg/h)	9.005e+004	9.005e+004
Std Ideal Liq Vol Flow	(m3/h)	99.37 *	99.37
Molar Enthalpy	(kJ/kgmole)	-7.581e+005	-7.581e+005
Molar Entropy	(kJ/kgmole-C)	665.4	665.4
Heat Flow	(kJ/h)	-1.909e+008	-1.909e+008
Liq Vol Flow @Std Cond	(m3/h)	99.37 *	99.37

Figure 4.50: Conditions of Mix 5 stream (Blend 8)

The following figures showing TBP curve (Figure 4.51) and Cuts Distribution (Figure 4.52).

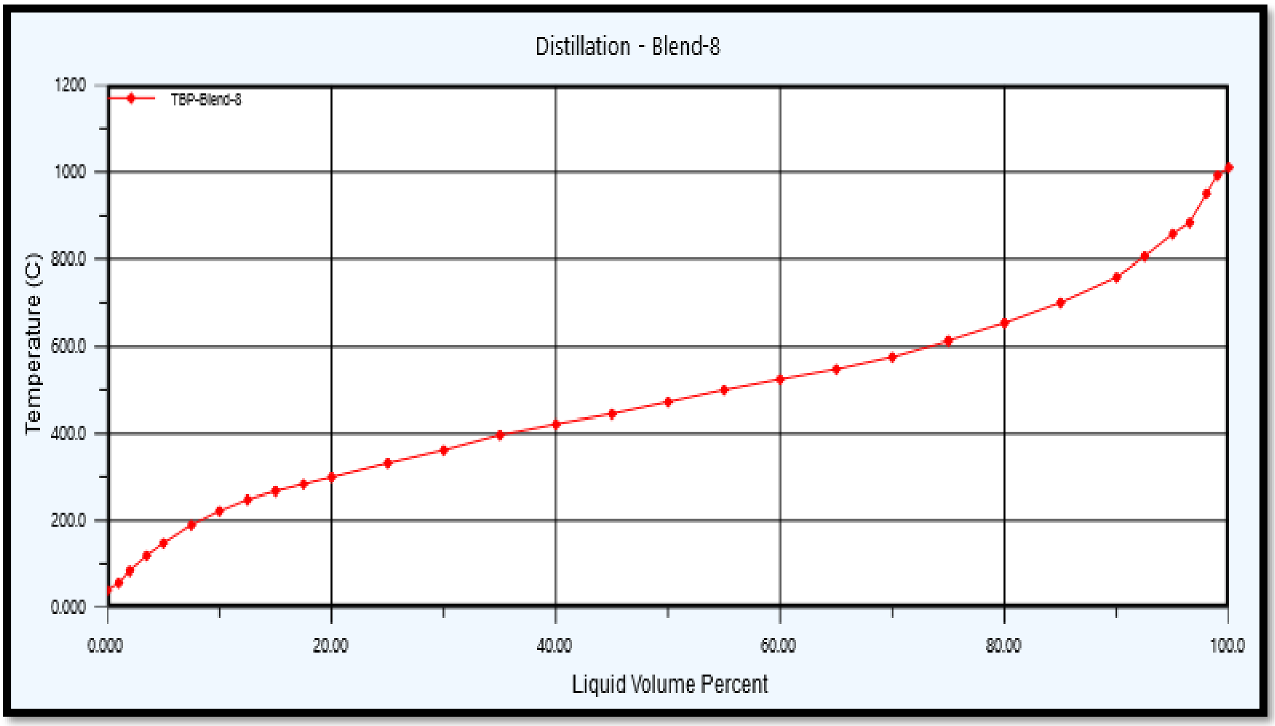


Figure 4.51: True Boiling Point Distillation Curve-Blend 8

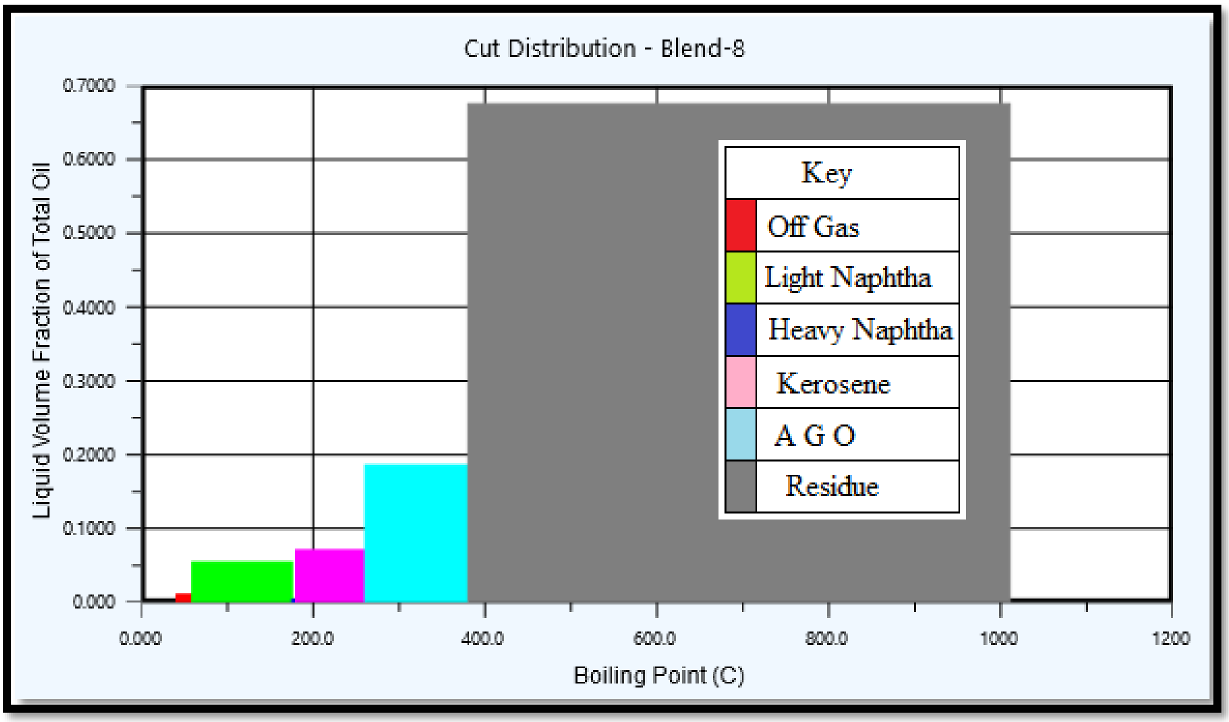


Figure 4.52: Cut Distribution-Blend 8

(Figure 4.53) showing the increasing of pressure during process

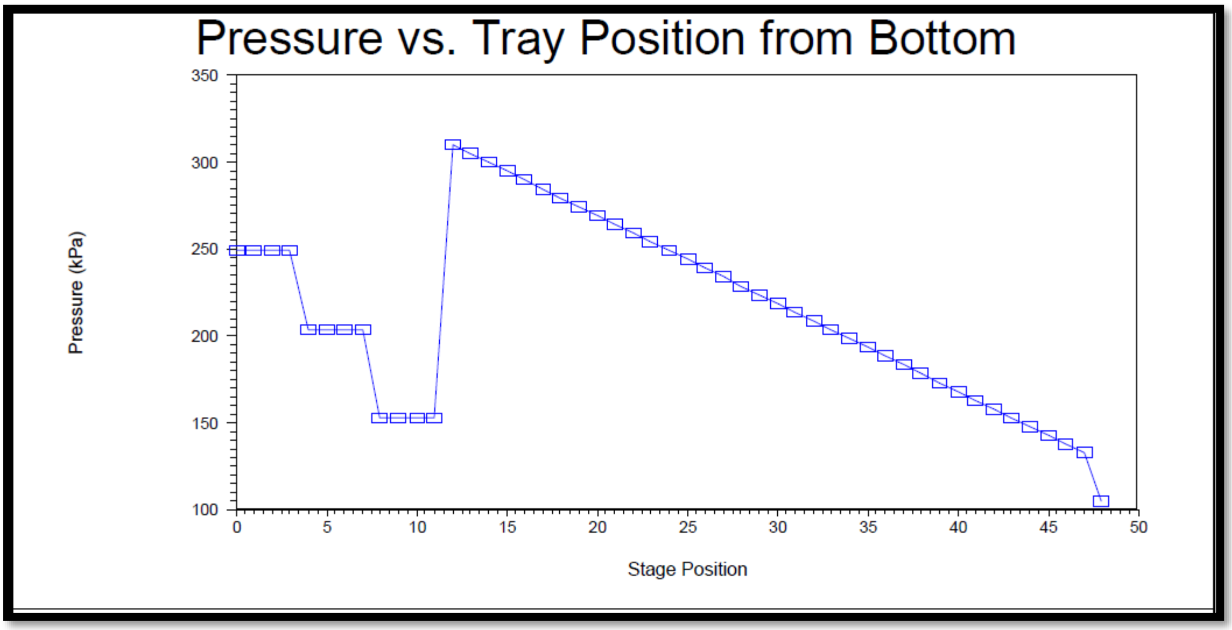


Figure 4.53: Pressure profile (Blend 8)

The change of temperature from tray to another shown in (figure 4.54)

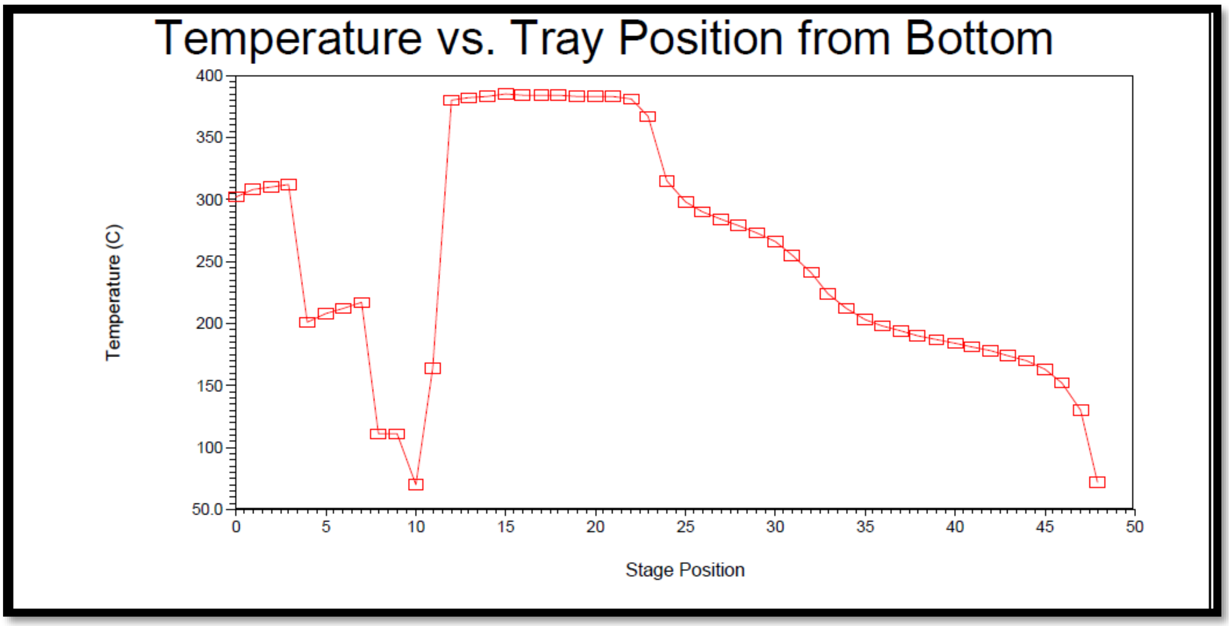


Figure 4.54: Temperature profile (Blend 8)

The quantity of liquid flow and Vapour flow shown in (figure 4.55)

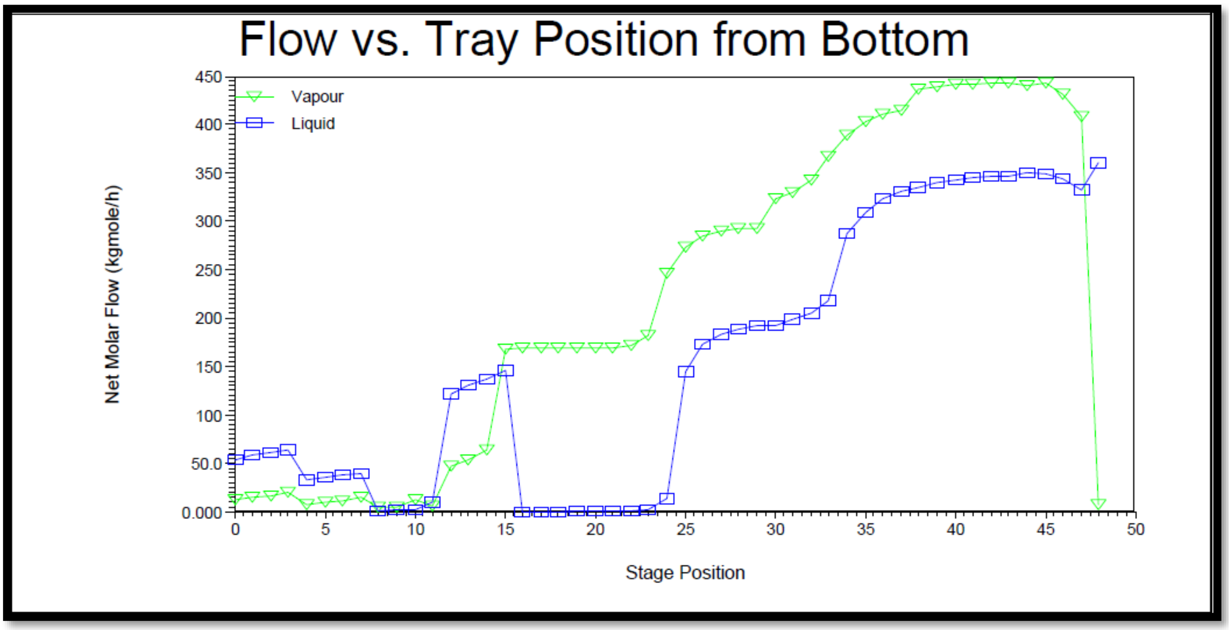


Figure 4.55: Net flow profile (Blend 8)

Column properties (Density and Molecular weight) shown in (figure 4.56)

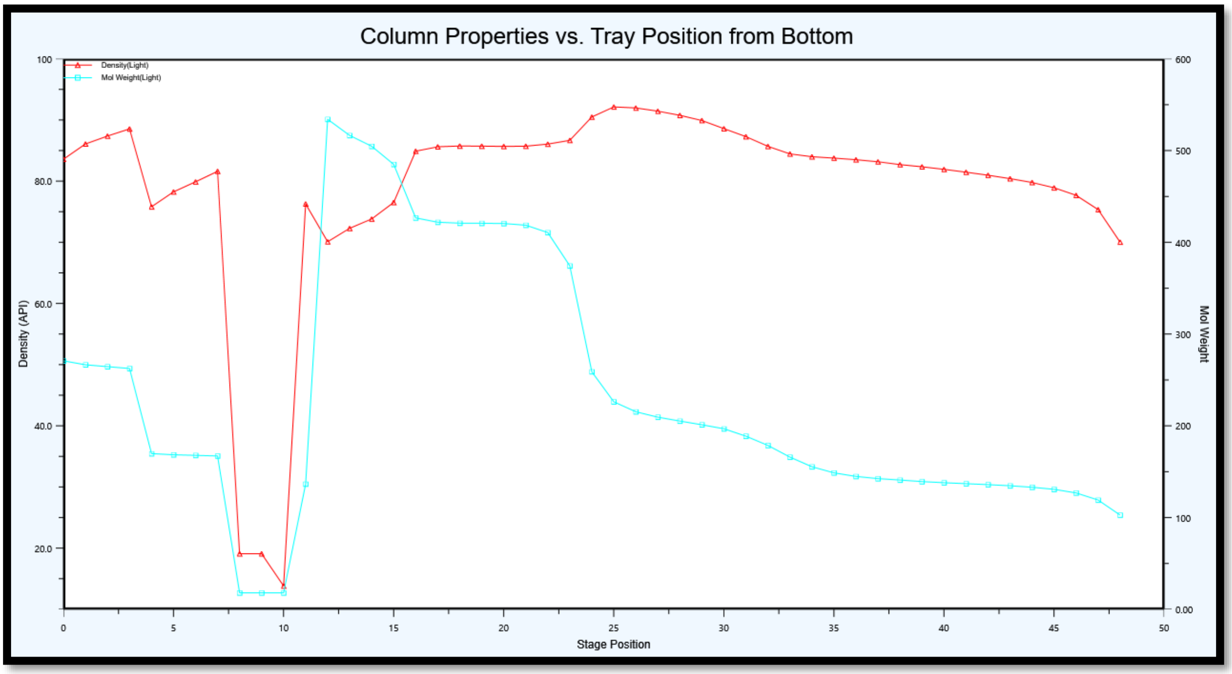


Figure 4.56: Correlation between density and molecular weight (Blend 8)

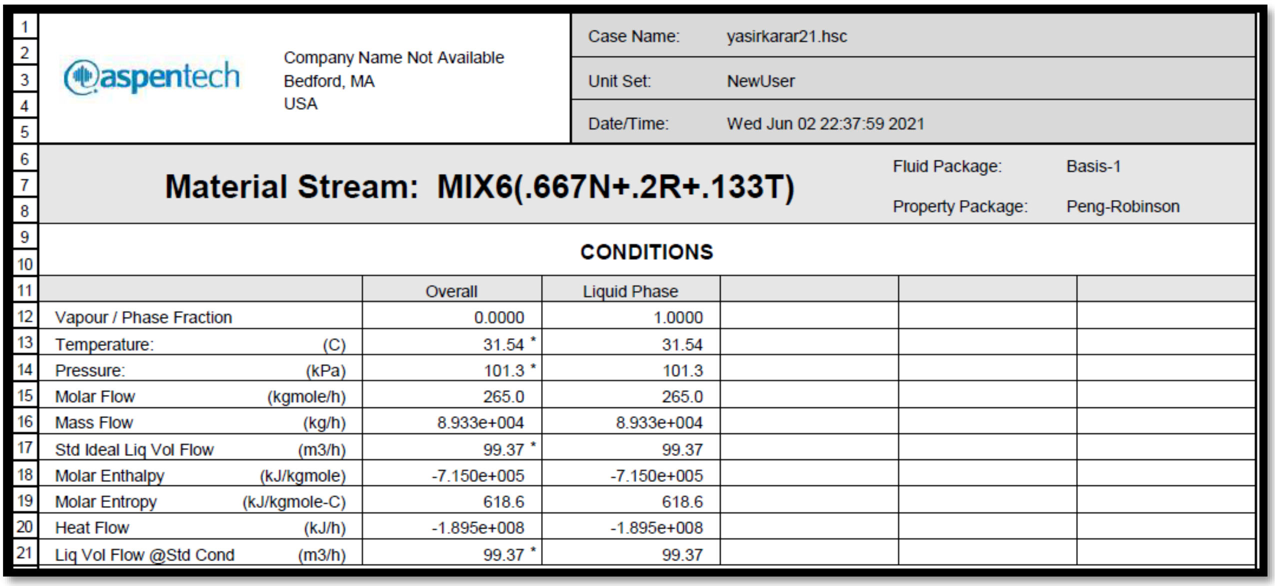
4.4.9 Case Study no.9 (Blend 9)

(53.33%Nile Blend+13.33%Rawat+33.33%Thargath) as defined Mix 6 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.10: The Oil Distributions -Blend 9

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	35.54	59	0.014
Light Naphtha	59	175	0.064
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.078
AGO	260	380	0.193
Residue	380	1005	0.647

(Figure 4.57) showing the Conditions of Mix 6 (blend 9) before entering the main distillation column.



Company Name Not Available
Bedford, MA
USA

Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:37:59 2021

Material Stream: MIX6(.667N+.2R+.133T)

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

		Overall	Liquid Phase
Vapour / Phase Fraction		0.0000	1.0000
Temperature:	(C)	31.54 *	31.54
Pressure:	(kPa)	101.3 *	101.3
Molar Flow	(kgmole/h)	265.0	265.0
Mass Flow	(kg/h)	8.933e+004	8.933e+004
Std Ideal Liq Vol Flow	(m3/h)	99.37 *	99.37
Molar Enthalpy	(kJ/kgmole)	-7.150e+005	-7.150e+005
Molar Entropy	(kJ/kgmole-C)	618.6	618.6
Heat Flow	(kJ/h)	-1.895e+008	-1.895e+008
Liq Vol Flow @Std Cond	(m3/h)	99.37 *	99.37

Figure 4.57: Conditions of Mix 6 stream (Blend 9)

The following figures showing TBP curve (Figure 4.58) and Cuts Distribution (Figure 4.59).

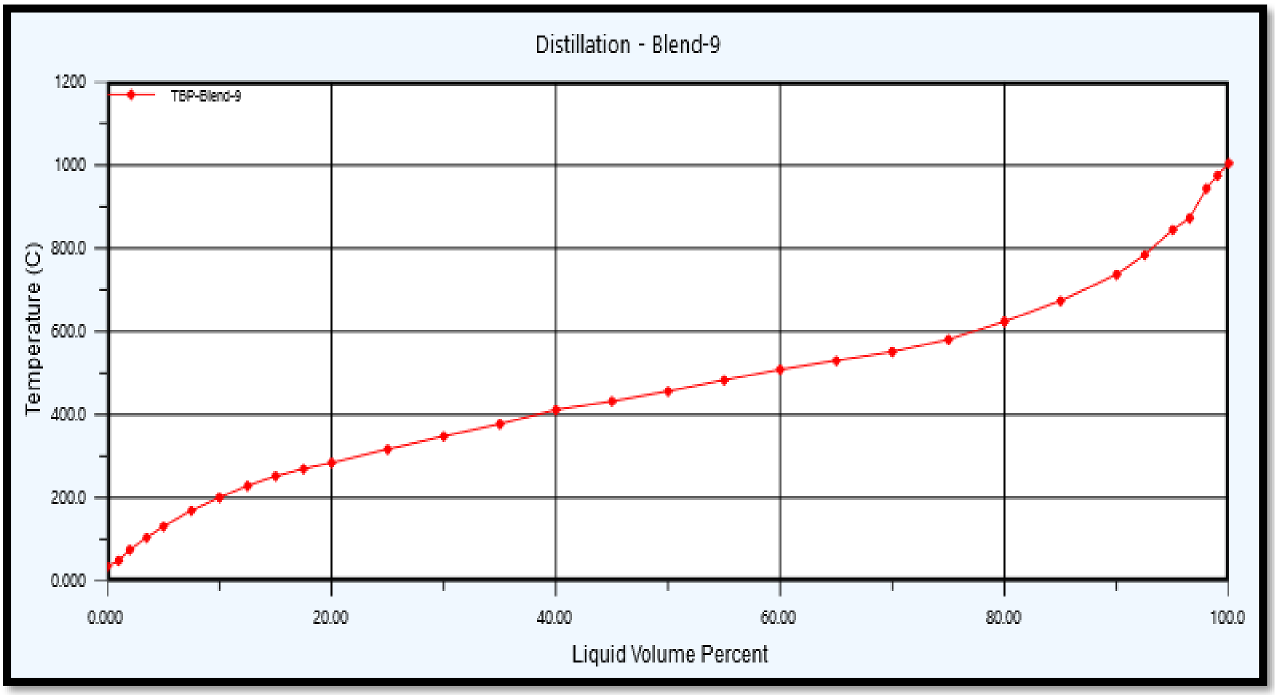


Figure 4.58: True Boiling Point Distillation Curve-Blend 9

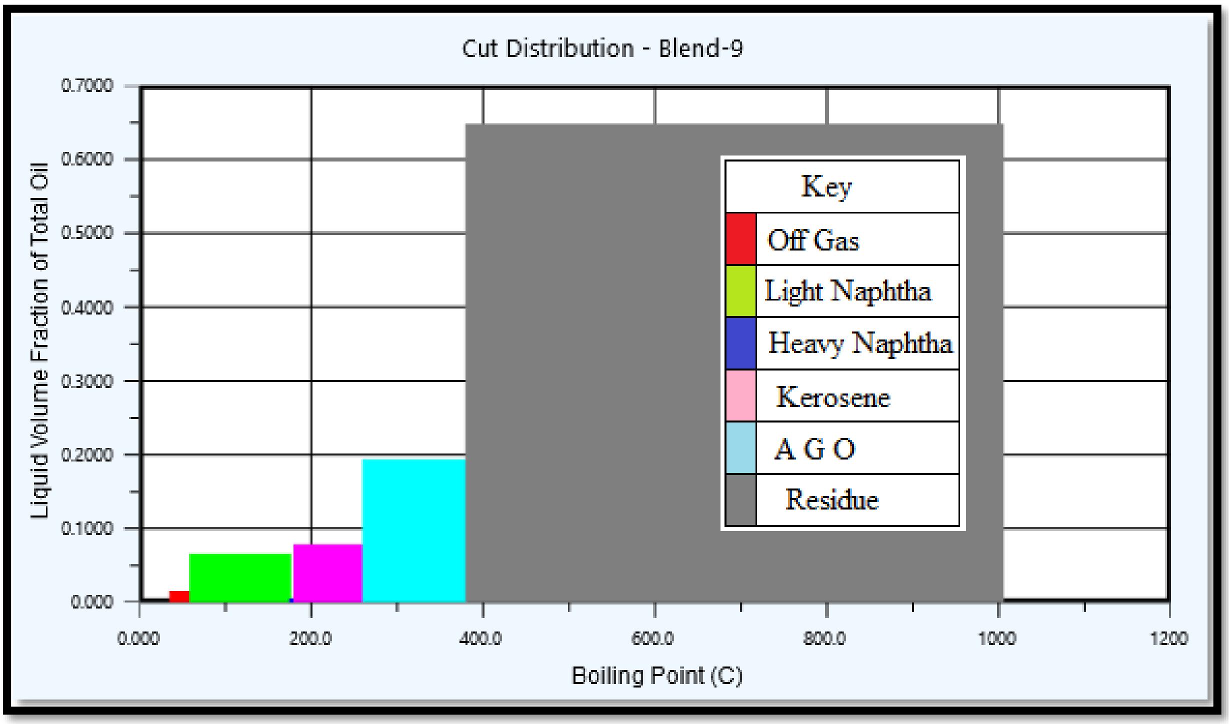


Figure 4.59: Cut Distribution-Blend 9

(Figure 4.60) showing the increasing of pressure during process

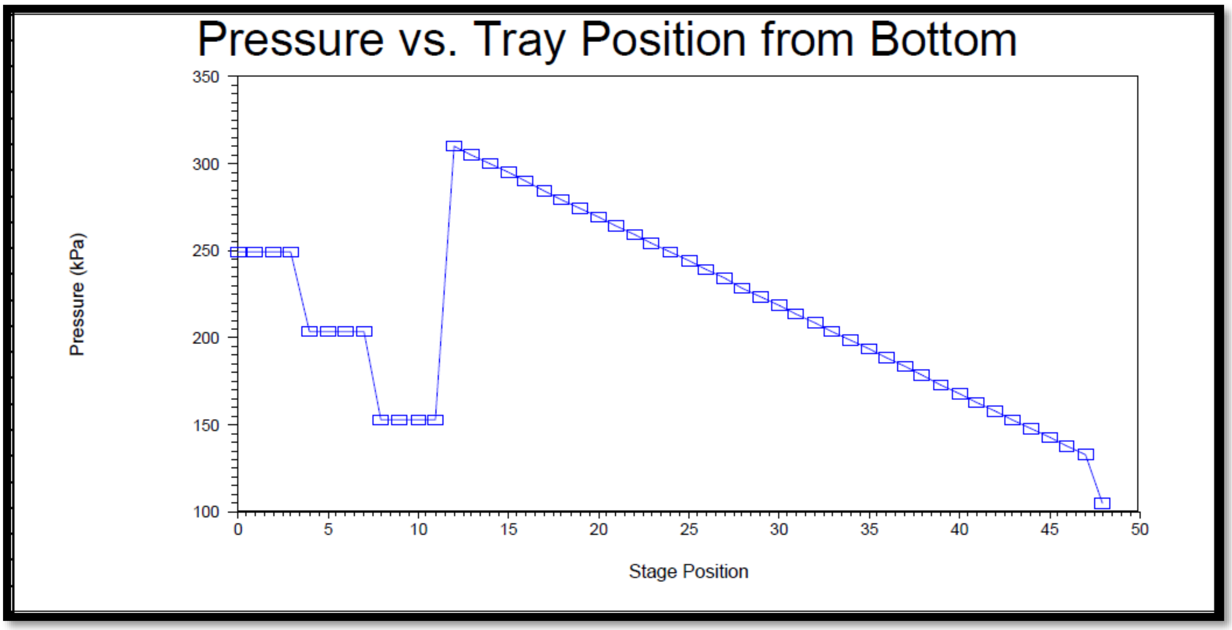


Figure 4.60: Pressure profile (Blend 9)

The change of temperature from tray to another shown in (figure 4.61)

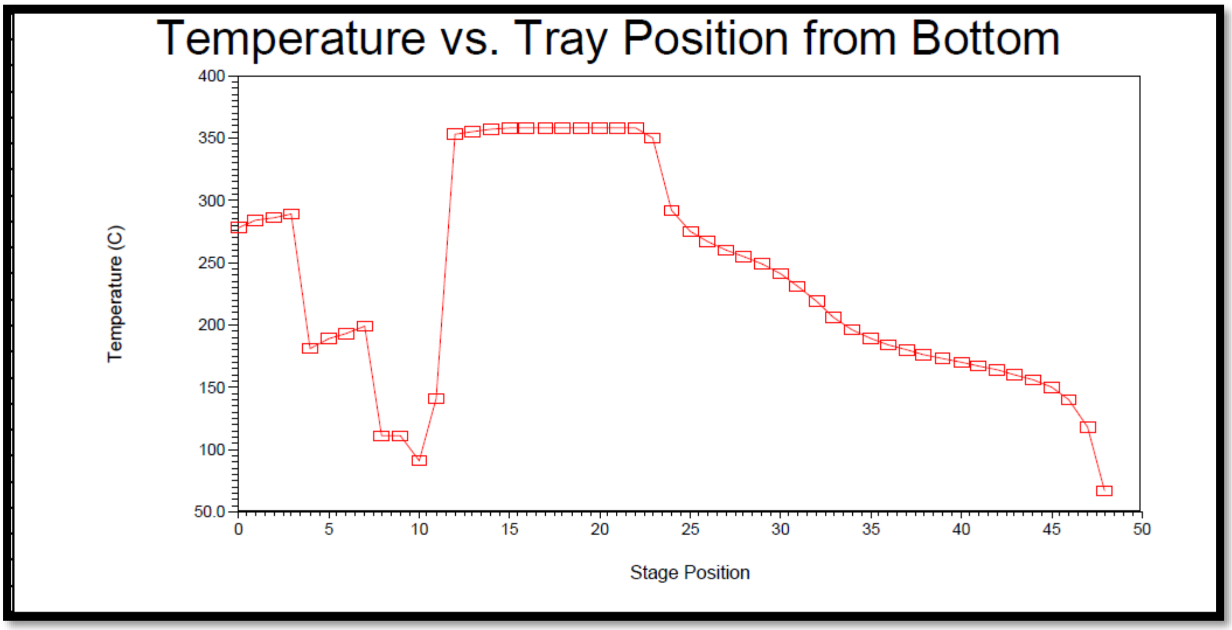


Figure 4.61: Temperature profile (Blend 9)

The quantity of liquid flow and Vapour flow shown in (figure 4.62)

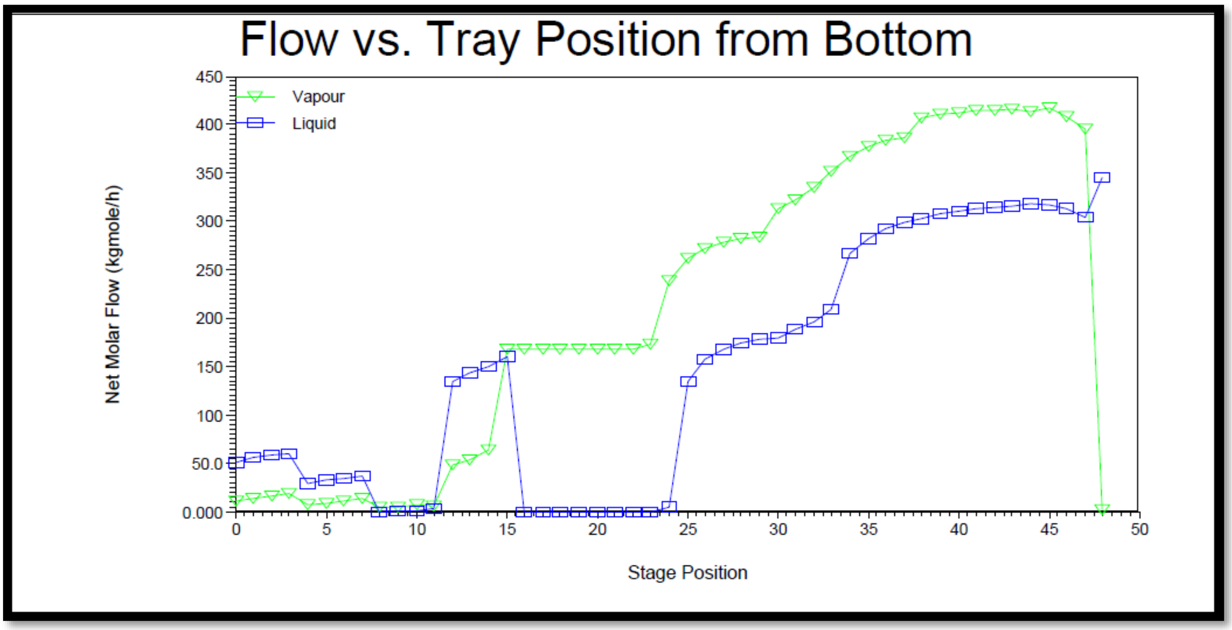


Figure 4.62: Net flow profile (Blend 9)

Column properties (Density and Molecular weight) shown in (figure 4.63)

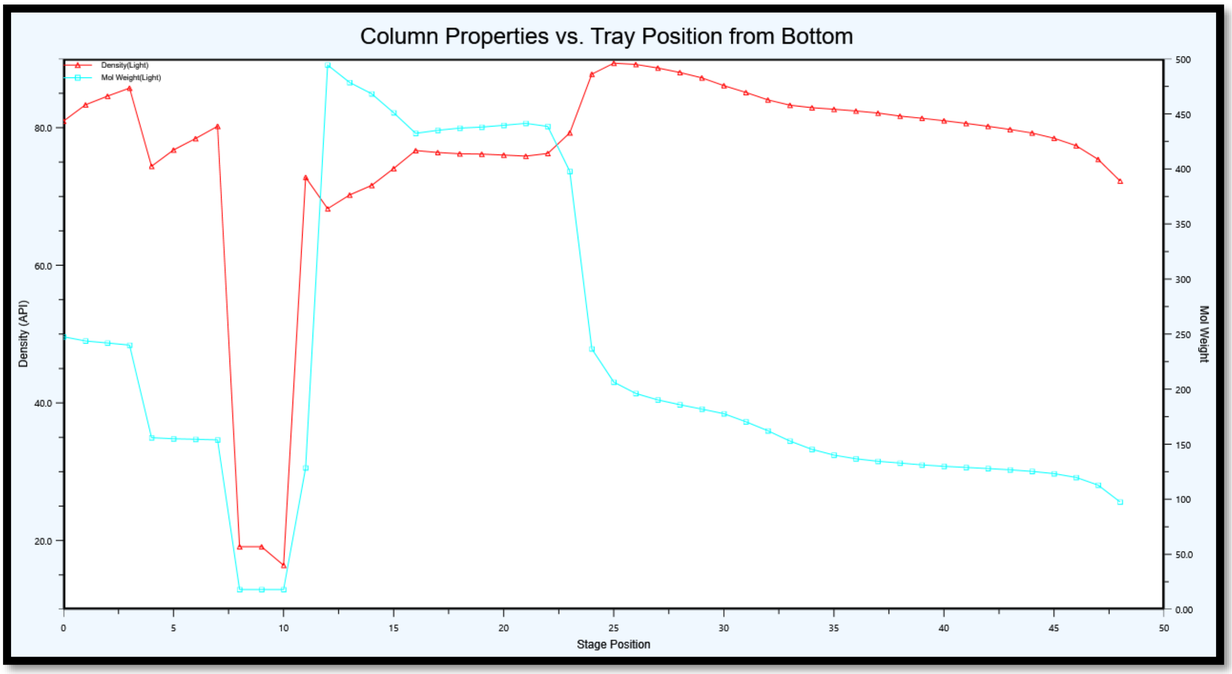


Figure 4.63: Correlation between density and molecular weight (Blend 9)

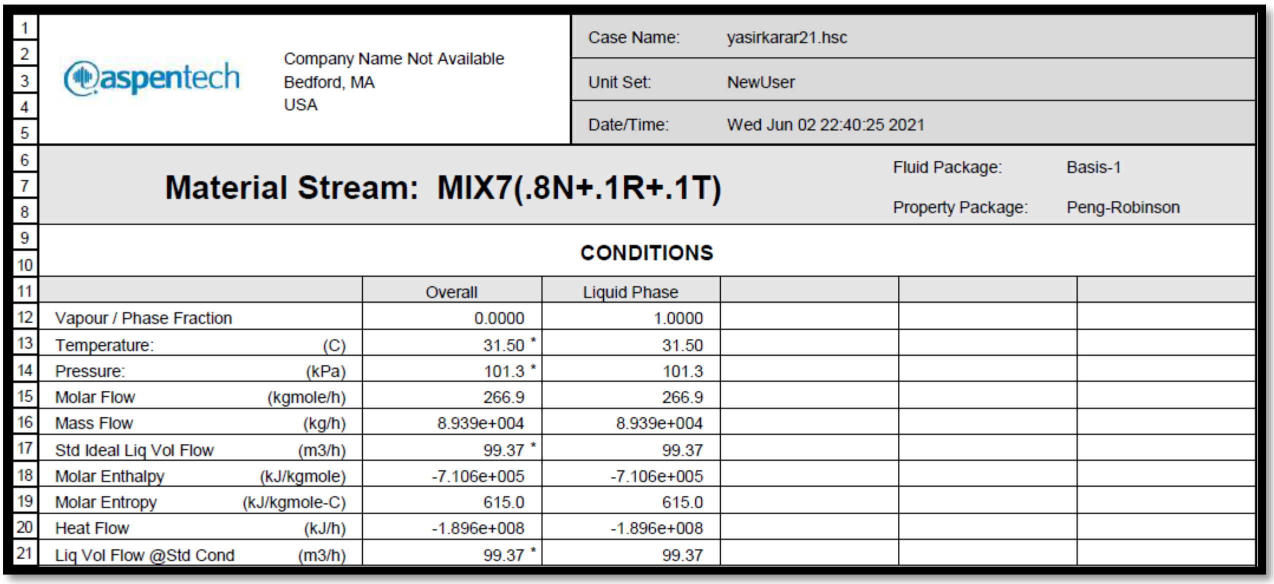
4.4.10 Case Study no.10 (Blend 10)

(80%Nile Blend+10%Rawat+10%Thargath) as defined Mix 7 crude was chosen, displaying the TBP curve and cutting the distribution, and the following results are obtained from Hysys.

Table 4.11: The Oil Distributions -Blend 10

Name	Initial T [°C]	Final T [°C]	Fraction Vol %
Off Gas	35.45	59	0.013
Light Naphtha	59	175	0.071
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.079
AGO	260	380	0.185
Residue	380	1010	0.647

(Figure 4.64) showing the Conditions of Mix 7 (blend 10) before entering the main distillation column.



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Company Name Not Available
Bedford, MA
USA

Case Name: yasirkarar21.hsc
Unit Set: NewUser
Date/Time: Wed Jun 02 22:40:25 2021

Material Stream: MIX7(.8N+.1R+.1T)

Fluid Package: Basis-1
Property Package: Peng-Robinson

CONDITIONS

		Overall	Liquid Phase
Vapour / Phase Fraction		0.0000	1.0000
Temperature:	(C)	31.50 *	31.50
Pressure:	(kPa)	101.3 *	101.3
Molar Flow	(kgmole/h)	266.9	266.9
Mass Flow	(kg/h)	8.939e+004	8.939e+004
Std Ideal Liq Vol Flow	(m3/h)	99.37 *	99.37
Molar Enthalpy	(kJ/kgmole)	-7.106e+005	-7.106e+005
Molar Entropy	(kJ/kgmole-C)	615.0	615.0
Heat Flow	(kJ/h)	-1.896e+008	-1.896e+008
Liq Vol Flow @Std Cond	(m3/h)	99.37 *	99.37

Figure 4.64: Conditions of Mix 7stream (Blend 10)

The following figures showing TBP curve (Figure 4.65) and Cuts Distribution (Figure 4.66).

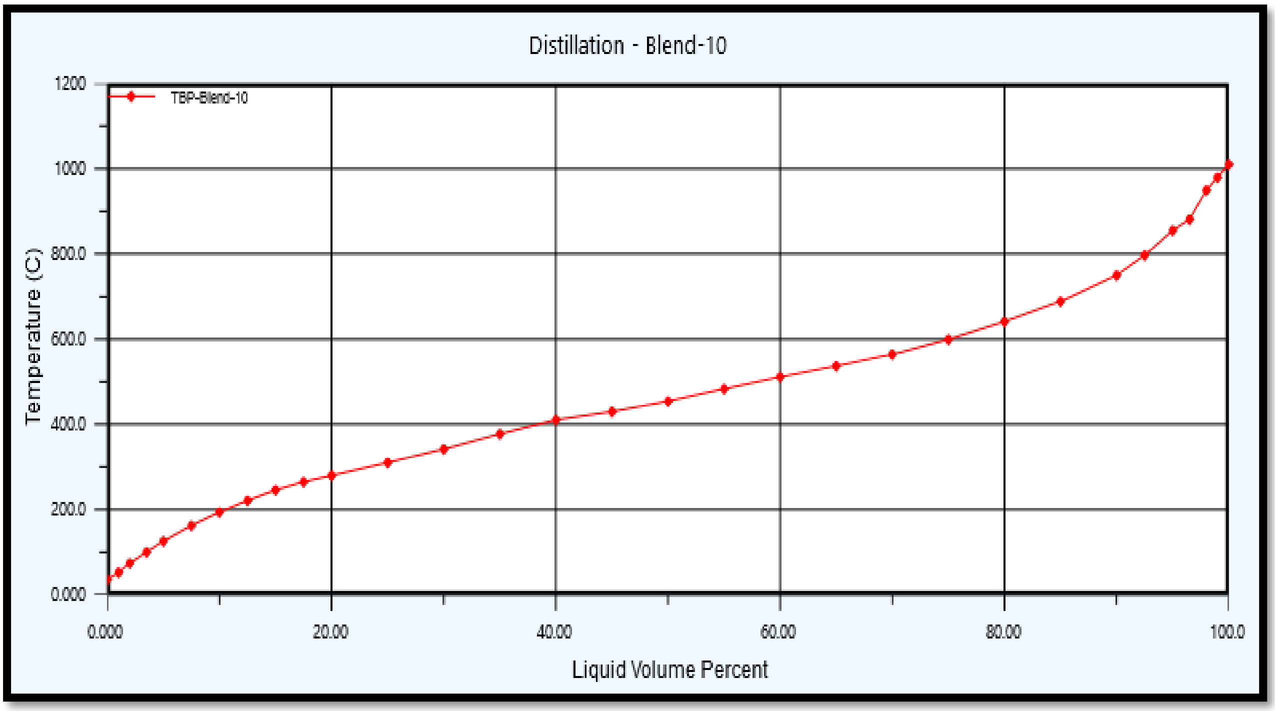

Figure 4.65: True Boiling Point Distillation Curve-Blend 10

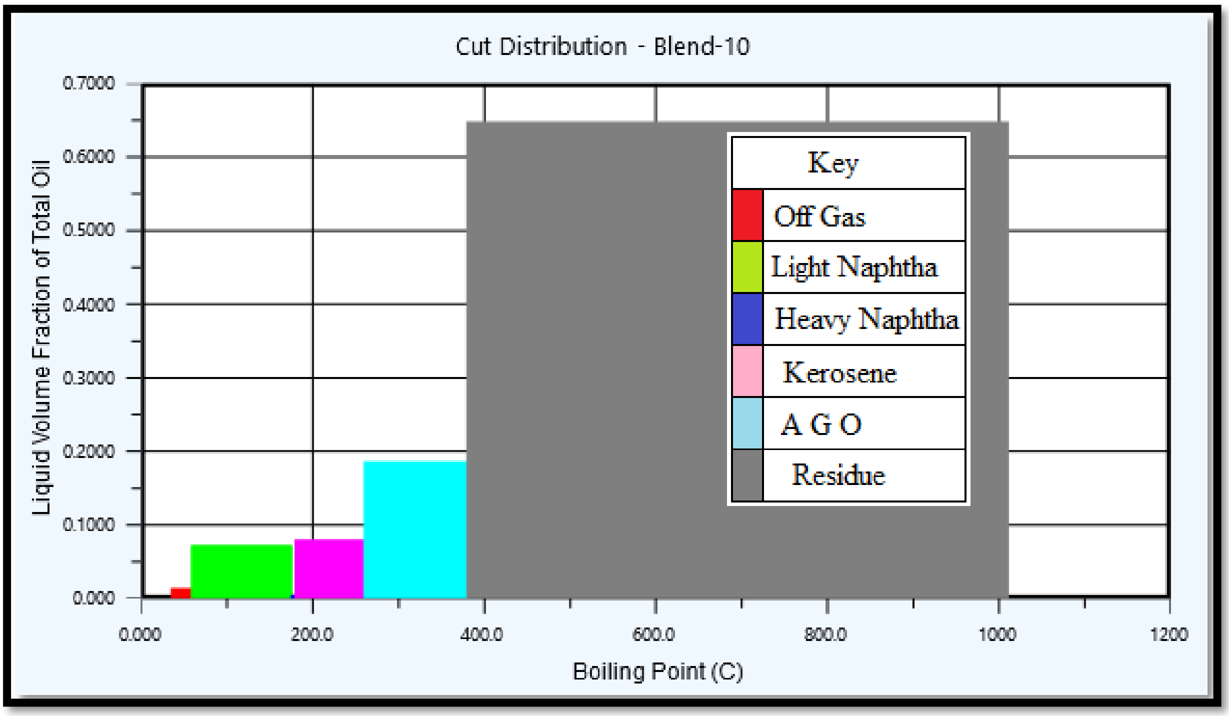

Figure 4.66: Cut Distribution-Blend 10

(Figure 4.67) showing the increasing of pressure during process.

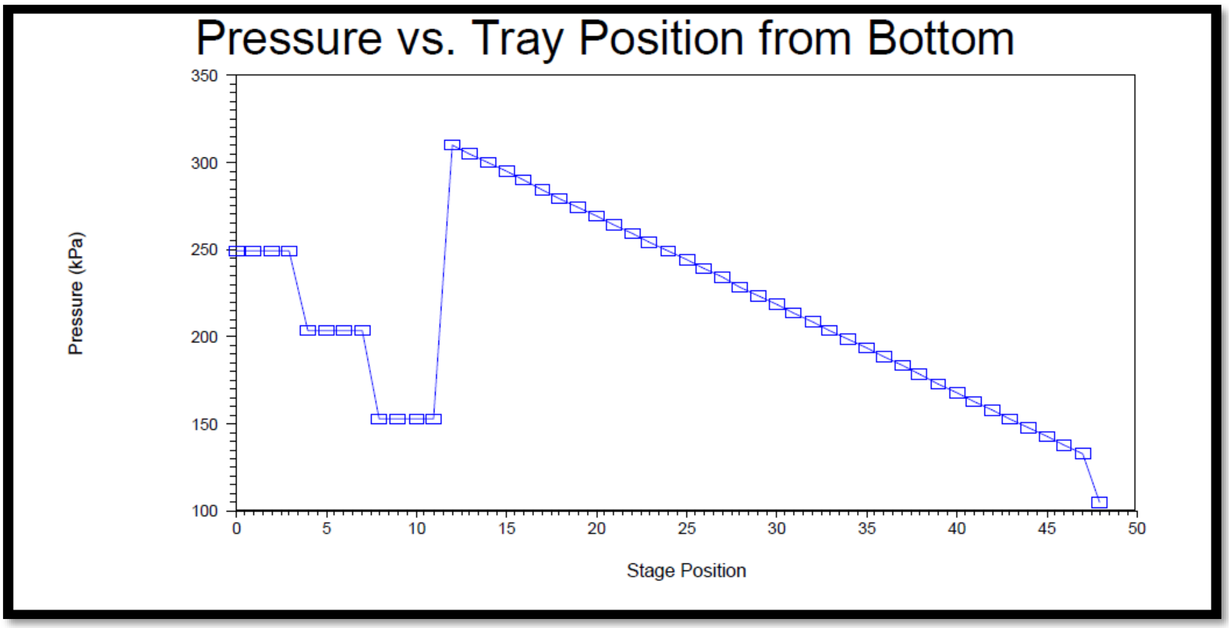

Figure 4.67: Pressure profile (Blend 10)

The change of temperature from tray to another shown in (figure 4.68).

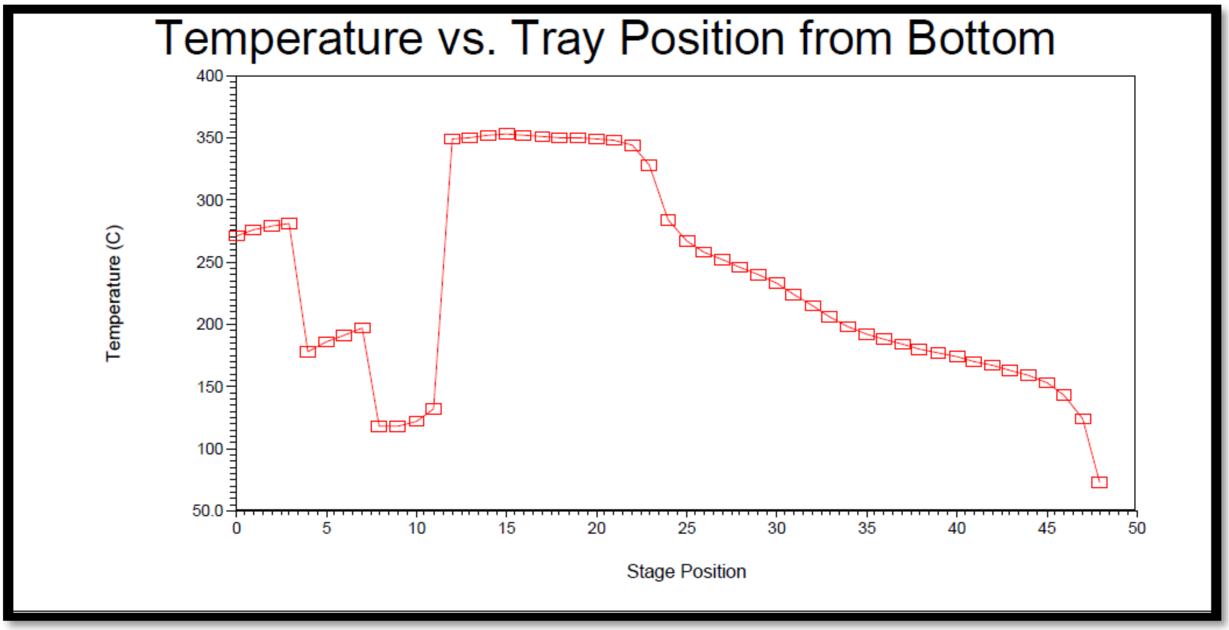

Figure 4.68: Temperature profile (Blend 10)

The quantity of liquid flow and Vapour flow shown in (figure 4.69).

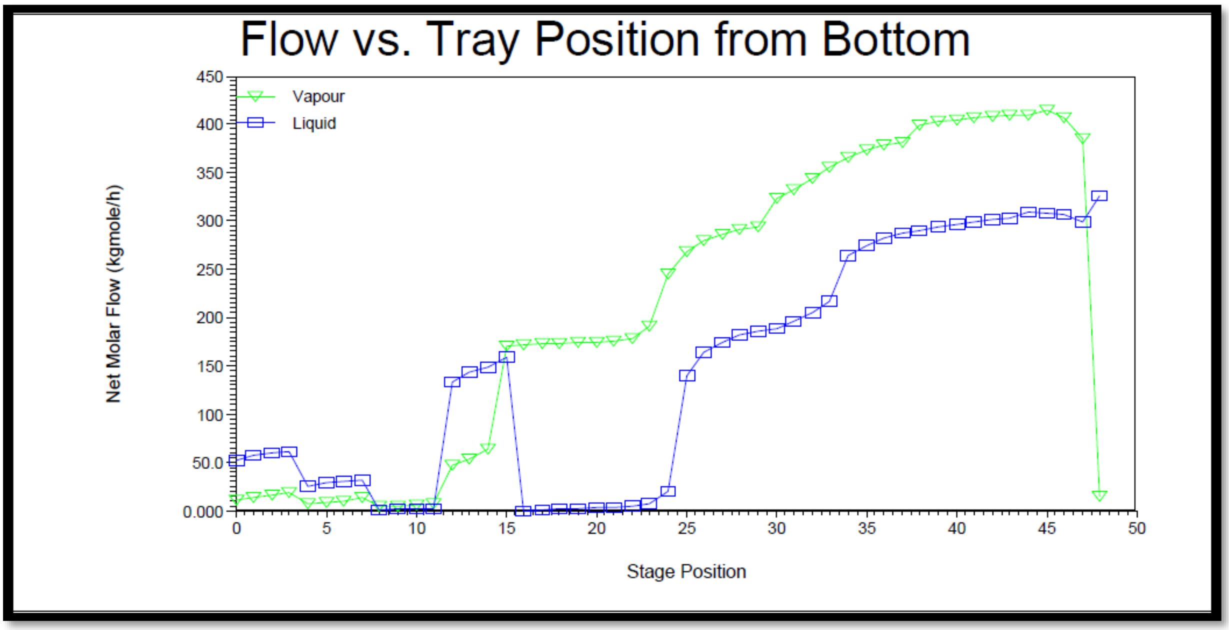

Figure 4.69: Net flow profile (Blend 10)

column properties (Density and Molecular weight) shown in (figure 4.70).

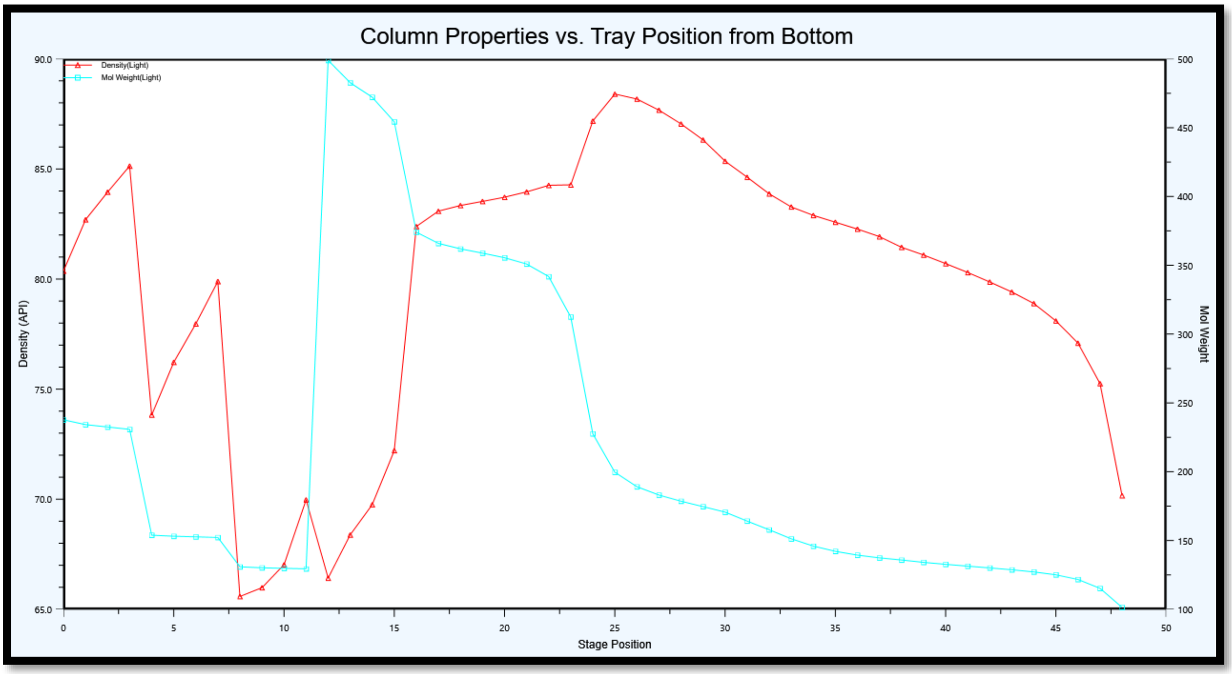

Figure 4.70: Correlation between density and molecular weight (Blend 10)

(Table 4.12) showing all the streams oil distribution by volume percentage, the flow rates for products of atmospheric distillation shown in (Table 4.13).

Table4.12: The oil distribution by volume percentage

Oil Distributions-Blend 1

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	34.4	59	0.015
Light Naphtha	59	175	0.079
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.084
AGO	260	380	0.180
Residue	380	1014	0.637

Oil Distributions-Blend 2

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	33.84	59	0.017
Light Naphtha	59	175	0.049
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.078
AGO	260	380	0.252
Residue	380	564.3	0.601

Oil Distributions-Blend 3

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	60	92.99	0.000
Light Naphtha	92.99	175	0.017
Heavy Naphtha	175	180	0.001
Kerosene	180	260	0.049
AGO	260	380	0.175
Residue	380	1027	0.758

Oil Distributions-Blend 4

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	36.98	59	0.013
Light Naphtha	59	175	0.061
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.075
AGO	260	380	0.191
Residue	380	1005	0.657

Oil Distributions-Blend 5

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	41.34	59	0.010
Light Naphtha	59	175	0.048
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.068
AGO	260	380	0.191
Residue	380	1007	0.680

Oil Distributions-Blend 6

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	37.43	59	0.012
Light Naphtha	59	175	0.063
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.076
AGO	260	380	0.186
Residue	380	1009	0.659

Oil Distributions-Blend 7

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	47.13	59	0.006
Light Naphtha	59	175	0.036
Heavy Naphtha	175	180	0.002
Kerosene	180	260	0.061
AGO	260	380	0.191
Residue	380	1010	0.703

Oil Distributions-Blend 8

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	40.35	59	0.011
Light Naphtha	59	175	0.055
Heavy Naphtha	175	180	0.003
Kerosene	180	260	0.071
AGO	260	380	0.186
Residue	380	1010	0.675

Oil Distributions-Blend 9

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	35.54	59	0.014
Light Naphtha	59	175	0.064
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.078
AGO	260	380	0.193
Residue	380	1005	0.647

Oil Distributions-Blend 10

Name	Initial T[°c]	Final T[°c]	Fraction
Off Gas	35.45	59	0.013
Light Naphtha	59	175	0.071
Heavy Naphtha	175	180	0.004
Kerosene	180	260	0.079
AGO	260	380	0.185
Residue	380	1010	0.647

Table 4.13: showed the crude oil Derivatives produce

Crude products	Blend1 Nile Blend 15000 bbl/Day 99.37 m^3/h	Blend2 Rawat 3000 bbl/Day 19.87 m^3/h	Blend3 Thargath 15000 bbl/Day 99.37 m^3/h	Blend4 Mix1 15000 bbl/Day 99.37 m^3/h	Blend5 Mix2 15000 bbl/Day 99.37 m^3/h	Blend6 Mix3 15000 bbl/Day 99.37 m^3/h	Blend7 Mix4 15000 bbl/Day 99.37 m^3/h	Blend8 Mix5 15000 bbl/Day 99.37 m^3/h	Blend9 Mix6 15000 bbl/Day 99.37 m^3/h	Blend10 Mix7 15000 bbl/Day 99.37 m^3/h
Reflux Ratio	5.609	2.230	22.22	4.453	8.434	5.931	6.923	7.083	5.955	5.159
Light Naphtha	7.255 m^3/h	3.00 m^3/h	1.8 m^3/h	7.867m^3/h	4.751 m^3/h	6.586 m^3/h	4.500 m^3/h	5.465 m^3/h	6.47 m^3/h	6.148 m^3/h
Reflux Rate	44.67 m^3/h	7.422 m^3/h	40.00 m^3/h	38.60 m^3/h	44.68 m^3/h	43.85 m^3/h	37.5 m^3/h	44.00 m^3/h	40.00 m^3/h	38.88 m^3/h
OFF GAS	0.7084 m^3/h	0.3287 m^3/h	0.00 m^3/h	0.8007m^3/h	0.5463 m^3/h	0.8070 m^3/h	0.9169 m^3/h	0.7474 m^3/h	0.2471 m^3/h	1.389 m^3/h
Residue	68.62 m^3/h	12.02 m^3/h	75.33 m^3/h	69.12 m^3/h	72.97 m^3/h	68.40 m^3/h	70.68 m^3/h	69.15 m^3/h	71.61 m^3/h	71.83 m^3/h
Waste Water Rate	0.8833 m^3/h	0.9682m^3/h	0.9827 m^3/h	0.8907m^3/h	0.9515 m^3/h	0.9297m^3/h	0.8931 m^3/h	0.9350 m^3/h	0.9693 m^3/h	0.8872m^3/h
Heavy Naphtha	0.05727m^3/h	0.07462m^3/h	0.09937m^3/h	0.030 m^3/h	0.03829m^3/h	0.0367m^3/h	0.05703m^3/h	0.04242m^3/h	0.02286m^3/h	0.3438m^3/h
Kerosene	4.832 m^3/h	0.585 m^3/h	4.869m^3/h	4.76 m^3/h	5.872 m^3/h	6.916 m^3/h	5.96 m^3/h	7.043 m^3/h	6.00 m^3/h	5.136 m^3/h
Atm Gas Oil	18.03 m^3/h	3.999 m^3/h	17.4 m^3/h	16.92 m^3/h	15.3 m^3/h	16.75 m^3/h	17.44 m^3/h	17.05 m^3/h	15.09 m^3/h	15.00 m^3/h

4.5 The Analysis of the Control Units and result of control by MATLAB

Transfer function identification (Gurashi ,2016)

Loop 1:

$G_c = K_c$ $\qquad$ $G_{V(S)} = \frac{2}{0.15S+1}$

$G_{P(S)} = \frac{0.8}{1.5S+1}$ $\qquad$ $G_{M(S)} = \frac{1}{0.05S+1}$

The block diagram:

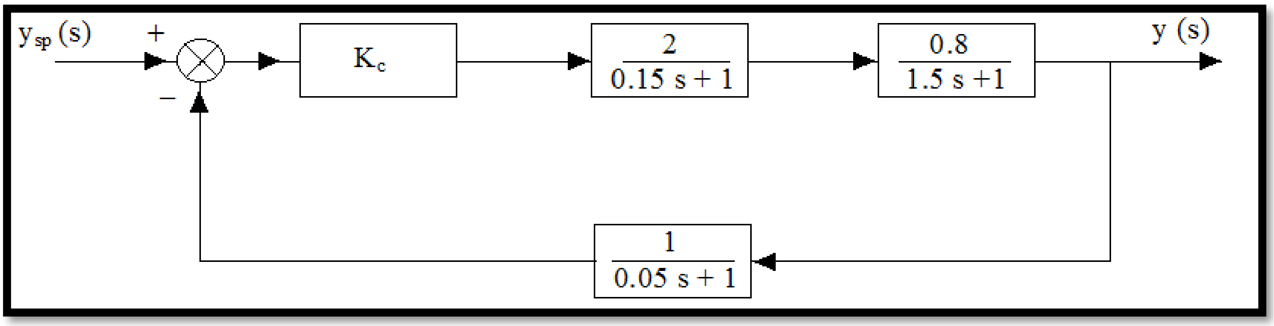

Figure 4.71: Feed Back Control Block Diagram

The overall transfer function:

$$G_{(S)} = \frac{\Pi_F}{1 \pm \Pi_l}$$

$$\Pi_F = K_C . \frac{2}{0.15s+1} . \frac{0.8}{1.5s+1} \Rightarrow \frac{1.6K_C}{(0.15s+1)(1.5s+1)}$$

$$\Pi_l = K_C . \frac{1.6}{(0.15s+1)(1.5s+1)} . \frac{1}{0.05s+1} \Rightarrow \frac{1.6K_C}{(0.15s+1)(1.5s+1)(0.05s+1)}$$

The characteristic equation : $\Pi_l + 1 = 0$

$$(0.15s+1)(1.5s+1)(0.05s+1) + 1.6K_C = 0$$

$$0.01125s^3 + 0.3075s^2 + 1.7s + 1 + 1.6K_C = 0$$

Using **Root-locus** and MATLAB, open loop transfer function were determined:

$$OLTF = \frac{1.6K_C}{0.01125s^3 + 0.3075s^2 + 1.7s + 1}$$

MATLAB format to get Root Locus:

```
>> num=[1.6];
>> den=[0.01125 0.3075 1.7 1];
>> rlocus(num,den)
>> grid
```

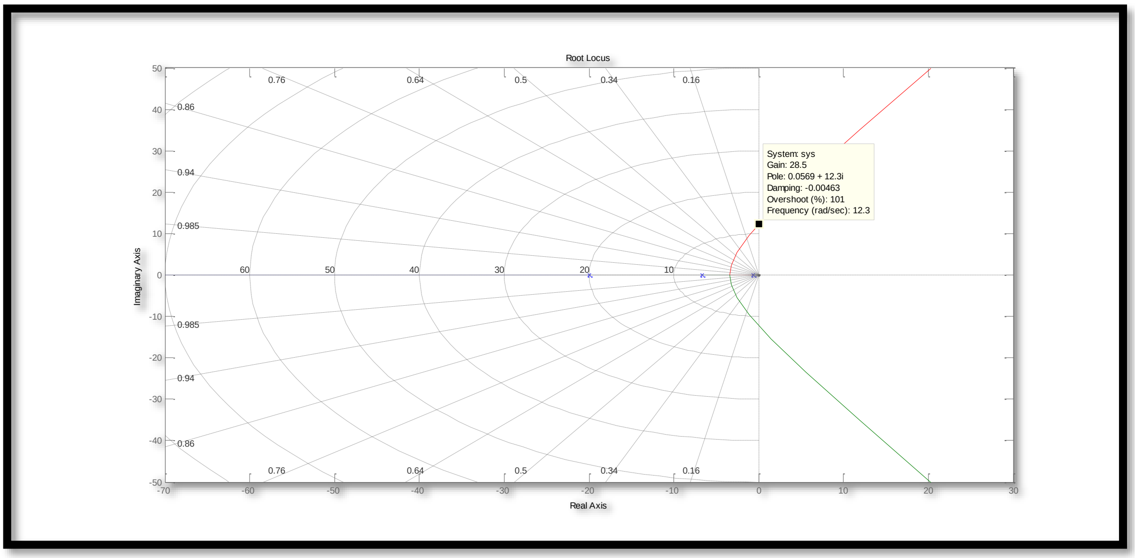

Figure 4.72: Root Locus plot of the system

K_u=28.5

w_{co}=12.3

$$P_U = \frac{2\Pi}{W_{CO}} = 0.511$$

Using **Bode plot** and MATLAB

$$OLTF = \frac{1.6K_{C1}}{0.01125s^3 + 0.3075s^2 + 1.7s + 1}$$

MATLAB format to get Bode:

```
>> num=[1.6];
>> den=[0.01125 0.3075 1.7 1];
>> bode(num,den)
>> grid
```

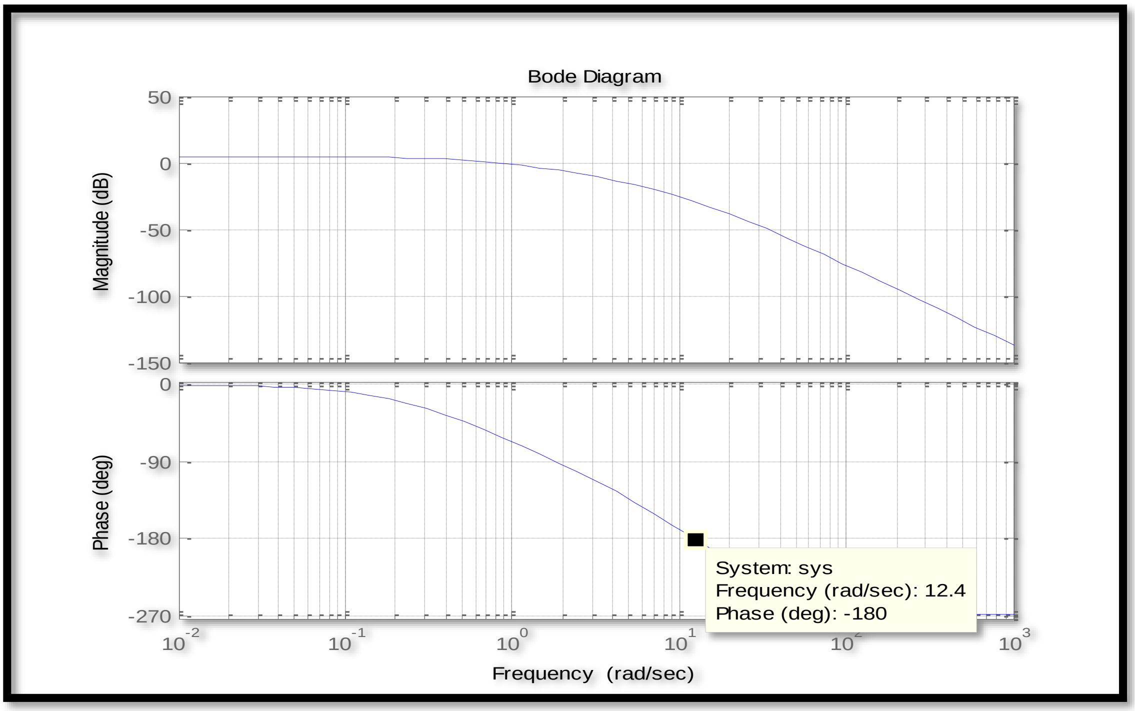

Figure 4.73: Bode plot for magnitude

W_{co}= 12.4 $P_u = 0.507$

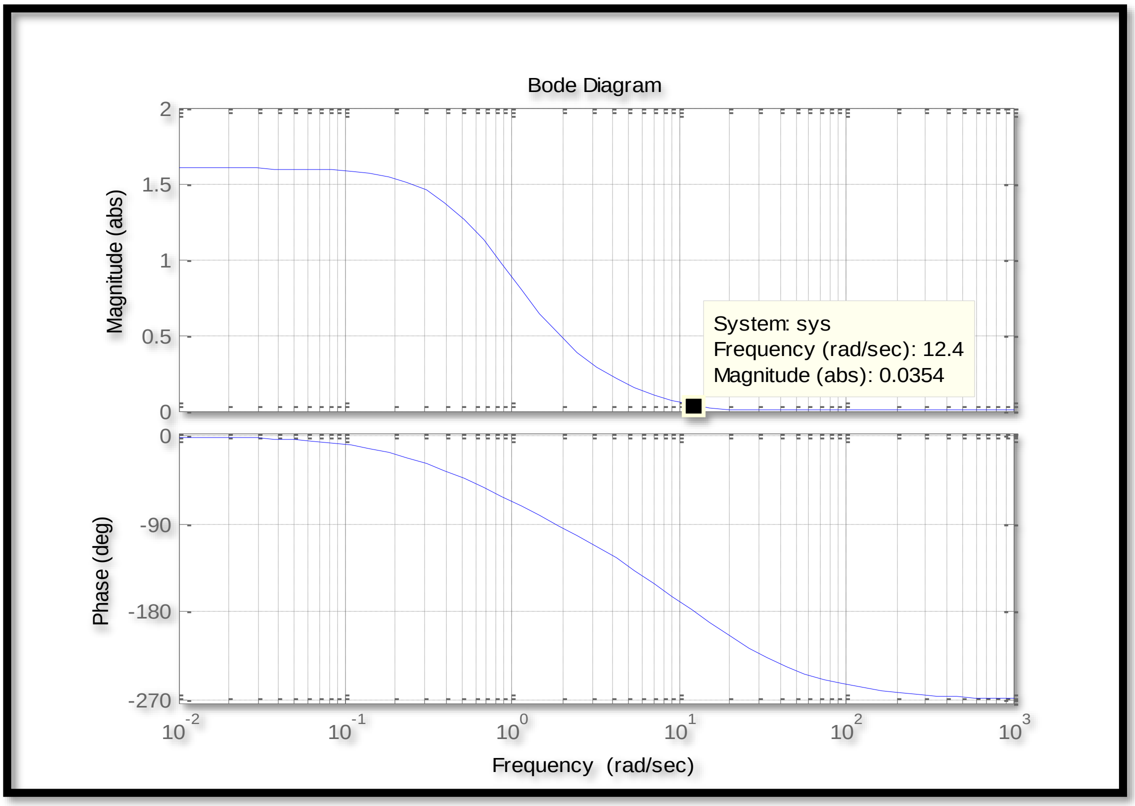

Figure 4.74: Bode plot for absolute

AR=0.0354

$$K_U = \frac{1}{AR} = 28.248$$

$$K_{U,average} = \frac{K_{U1} + K_{U2}}{2} = 28.37 \qquad P_{U,average} = \frac{P_{U1} + P_{U2}}{2} = 0.509$$

The following is **Ziegler-Nichols** tuning table:

Table 4.14: Ziegler-Nichols tuning

Type of controller	**K_C**	**τ_I**	**τ_D**
P	0.5 K_u = 14.185	**-**	**-**
PI	0.45 K_u = 12.76	$\frac{P_U}{1.2} = 0.4242$	**-**
PID	0.6 K_u = 17.022	$\frac{P_U}{2} = 0.2545$	$\frac{P_U}{8} = 0.06363$

Overshoot for P; PI; PID-controller, the closed-loop transfer function:

$$G_{(S)} = \frac{\dfrac{1.6K_C}{(0.15s+1)(1.5s+1)}}{1 + \dfrac{1.6K_C}{(0.15s+1)(1.5s+1)(0.05s+1)}} = \frac{1.6K_C(0.05s+1)}{0.01125s^3 + 0.3075s^2 + 1.7s + 1 + 1.6K_C}$$

4.5.1 Proportional Controller Overshoot (P):

MATLAB format to get the step response for Proportional Controller

```
>> num=[1.135 22.696];
>> den=[0.01125 0.3075 1.7 23.696];
>> step(num,den)
>> grid
```

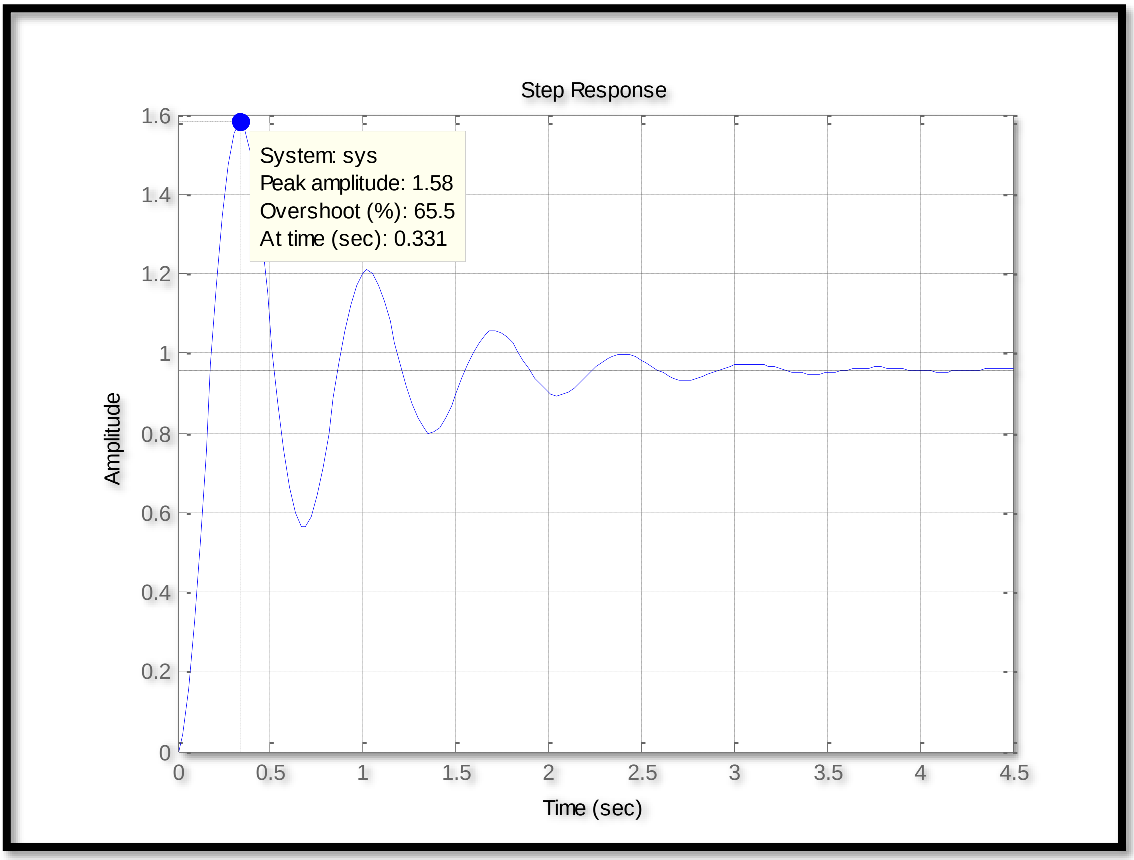

Figure 4.75: Unit step response for P – controller

4.5.2 Proportional Integral Controller Overshoot (PI):

MATLAB format to get the step response for Proportional Integral Controller

```
>> num=[1.021 20.416];
>> den=[0.01125 0.3075 1.7 21.416];
>> step(num,den)
>> grid
```

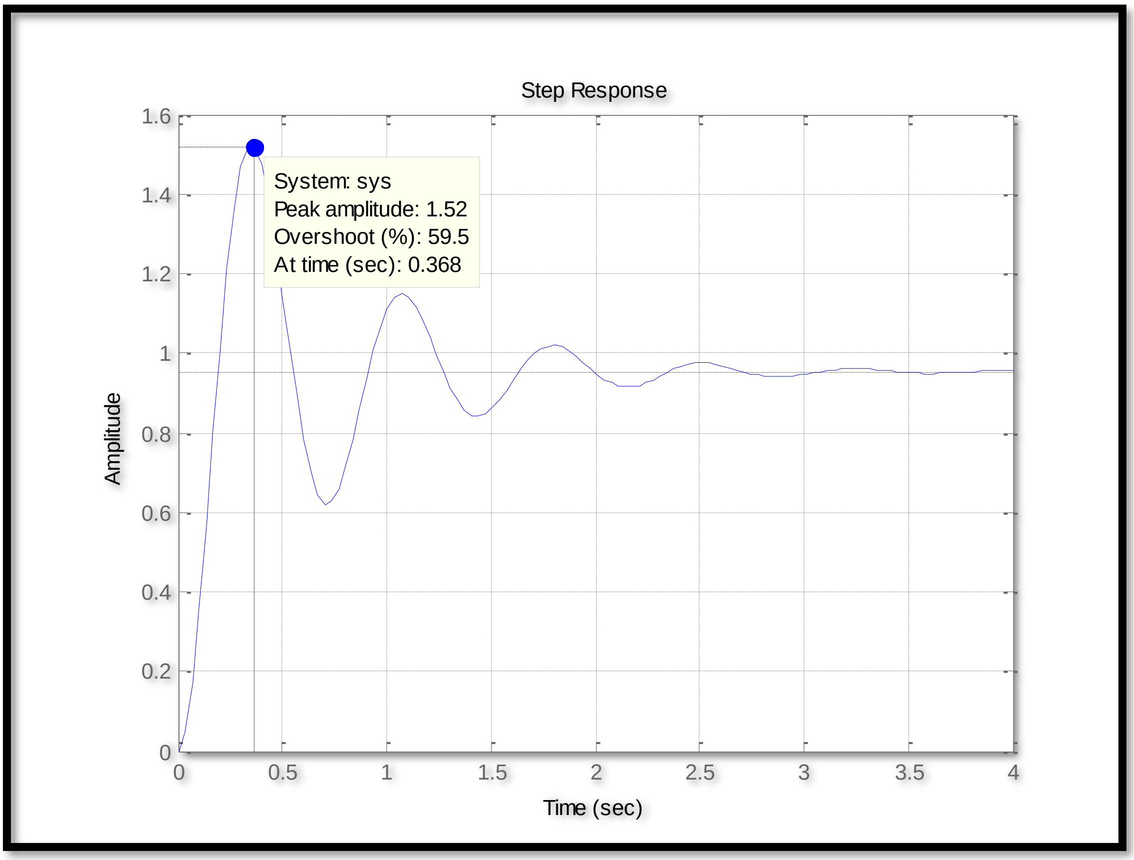

Figure 4.76: Unit step response for PI – controller

4.5.3 Proportional Integral Derivative Controller Overshoot (PID):

MATLAB format to get the step response for Proportional integral derivative Controller

```
>> num=[1.361 27.235];
>> den=[0.01125 0.3075 1.7 28.235];
>> step(num,den)
>> grid
```

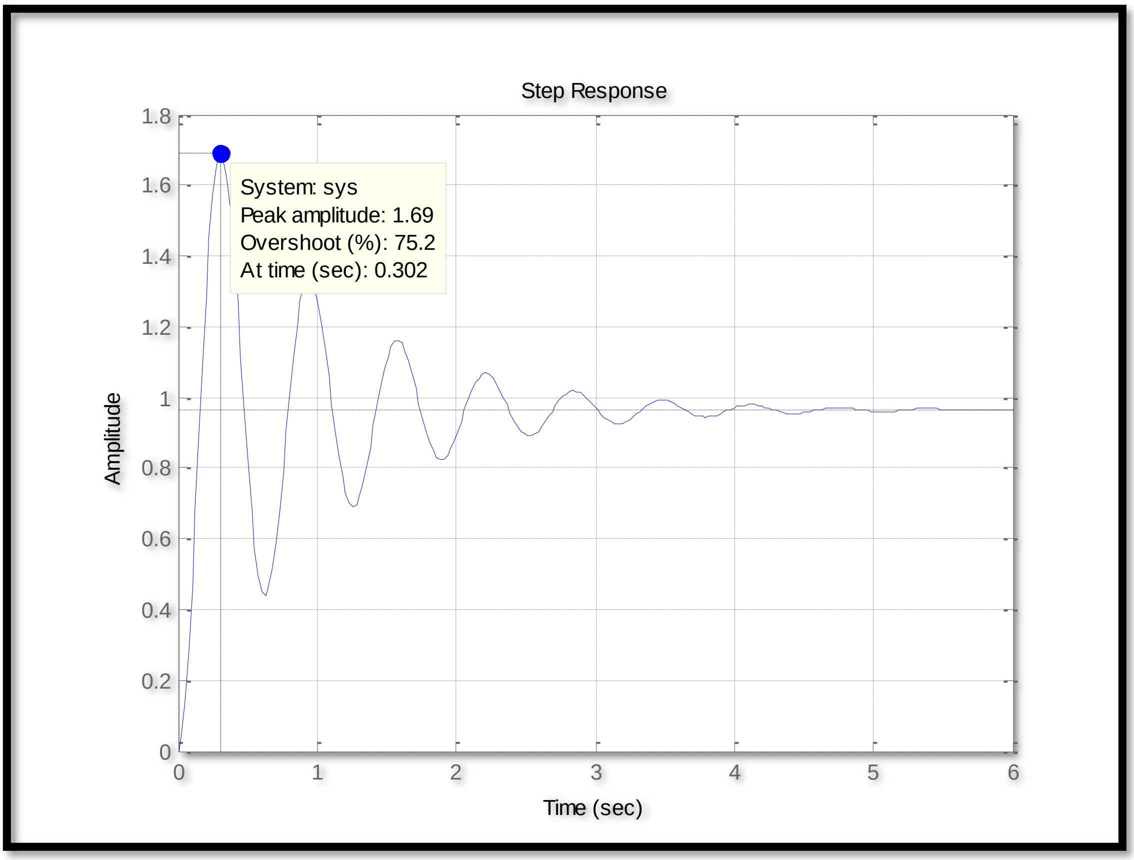

Figure 4.77: Unit step response for PID – controller

Table 4.15: Comparison Between (P, PI, PID) Controller

Type of controller	Overshoot(%)	Gain Kc	Rise Time(sec)
P	65.5	14.185	0.121
PI	59.5	12.76	0.130
PID	75.2	17.022	0.108

Overshoot for three type P, PI and PID controller as shown in (Figure 4.78)

MATLAB format to get the Comparison between three controllers (P, PI and PID)

```
>>num1= [1.135 22.696];
>>den1= [0.01125 0.3075 1.7 23.696];
>>num2= [1.021 20.416];
>>den2= [0.01125 0.3075 1.7 21.416];
>>num3= [1.361 27.235];
>>den3= [0.01125 0.3075 1.7 28.235];
>>hold on
>>step (num1, den1)
>>step (num2, den2)
>>step (num3, den3)
>>grid
>>legend('P','PI','PID')
>>hold off
```

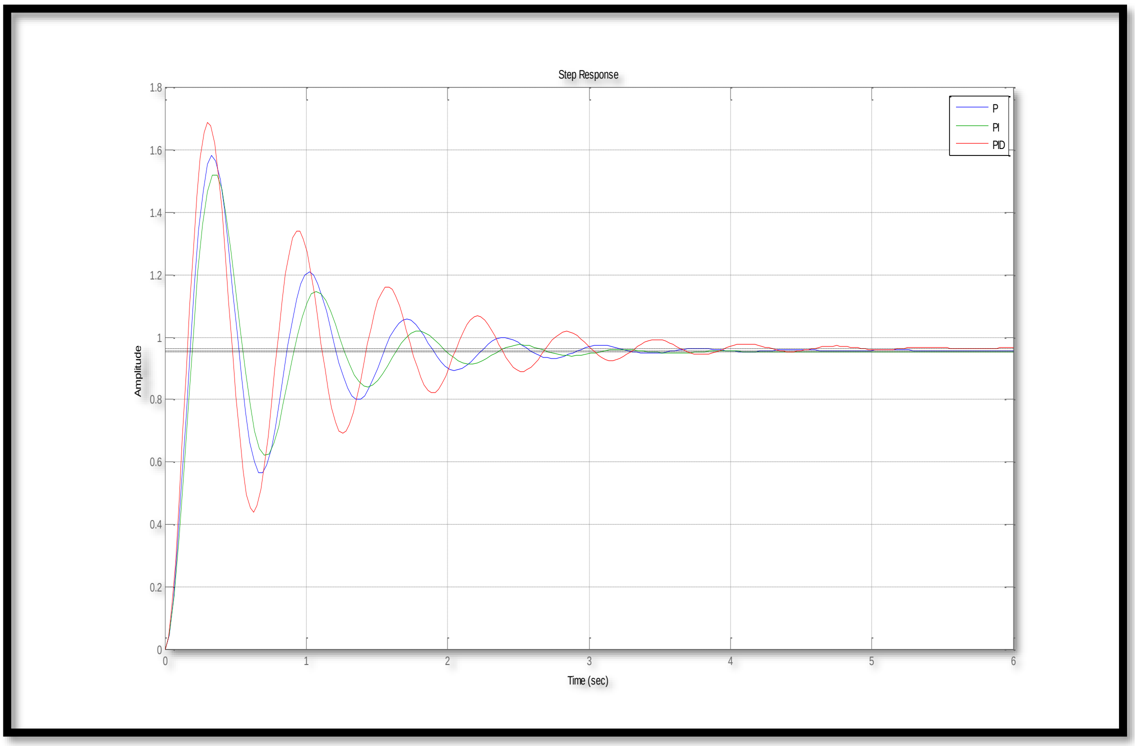

Figure 4.78: Unit step response for the (P, PI, PID) Controller

Proportional integral controller was selected to use for the system for Controlling of the temperature of vacuum furnace by control of the fuel flowrate, it provides a minimum overshoot of the P and PID controllers in changing the step and making the system stable.

4.6 Vacuum Distillation Unit Result

4.6.1 Case Study no.1

Residue of the stream Nile Blend was chosen for entering in the vacuum distillation tower to convert it to light products.

The change of temperature from tray to another shown in (figure 4.79).

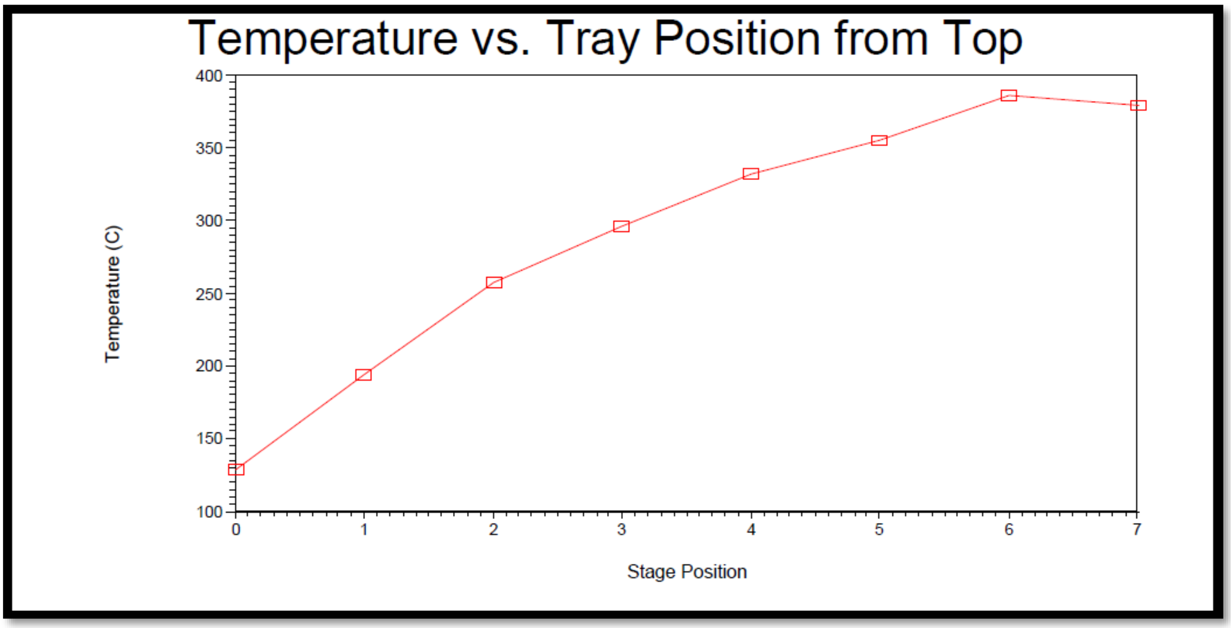

Figure 4.79: Temperature profile for Nile blend (vacuum distillation)

(Figure 4.80) showing the increasing of pressure during process

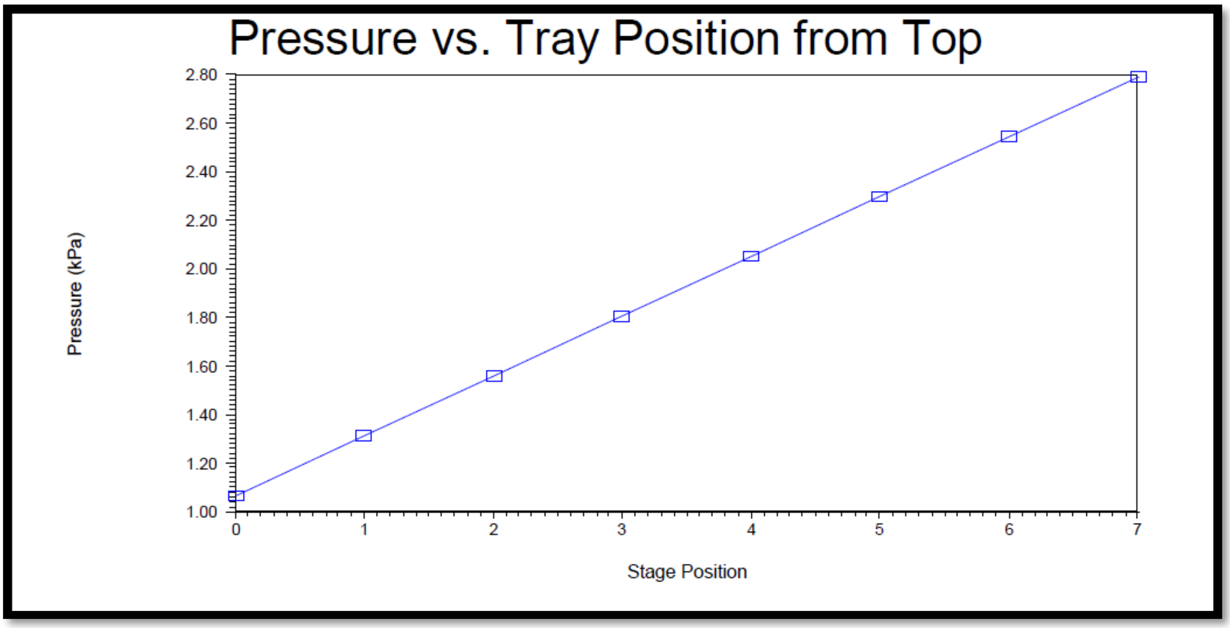

Figure 4.80: Pressure profile for Nile blend (vacuum distillation)

The quantity of liquid flow and Vapour flow shown in (figure 4.81).

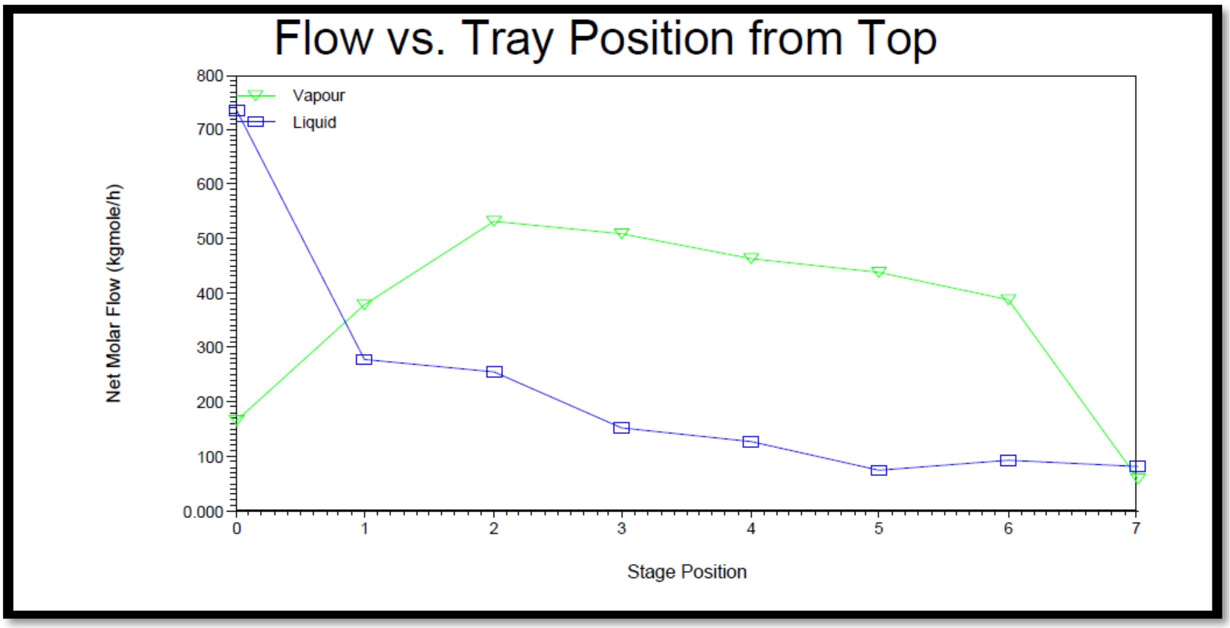

Figure 4.81: Net flow profile for Nile blend Figure (vacuum distillation)

column properties (Density and Molecular weight) shown in (figure 4.82).

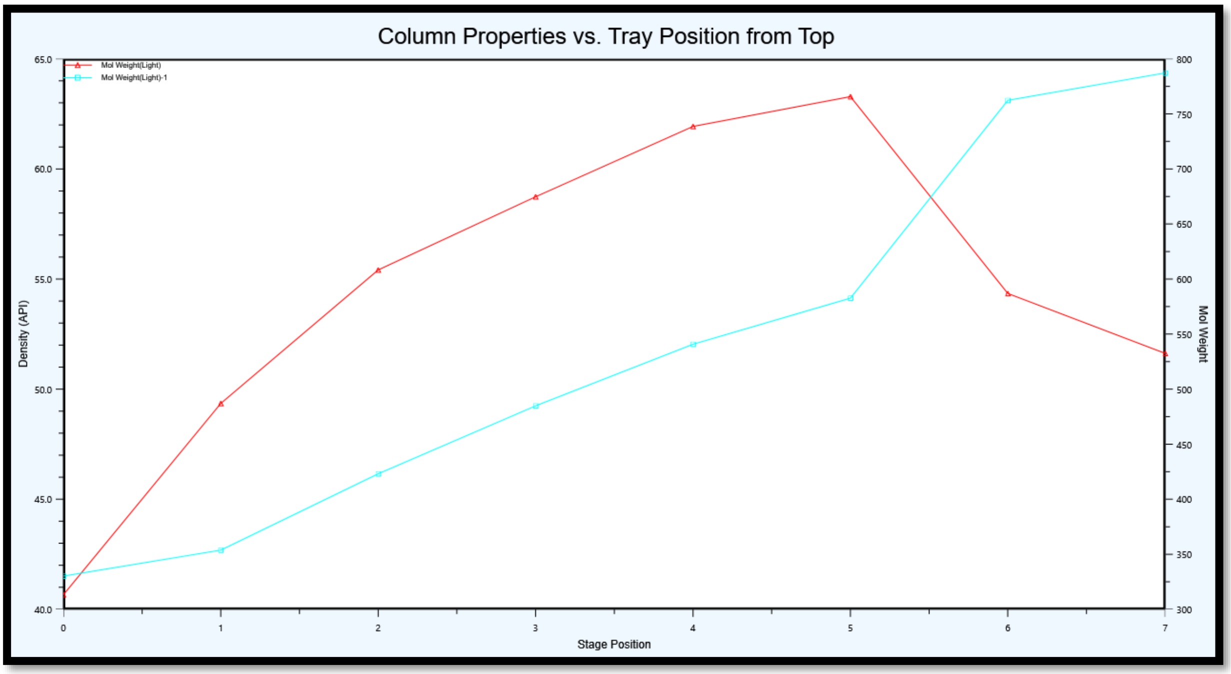

Figure 4.82: Correlation between density and molecular weight for Nile blend (vacuum distillation)

4.6.2 Case Study no.2

Residue of the stream Mix 1 was chosen for entering in the vacuum distillation tower to convert it to light products.

The change of temperature from tray to another shown in (figure 4.83).

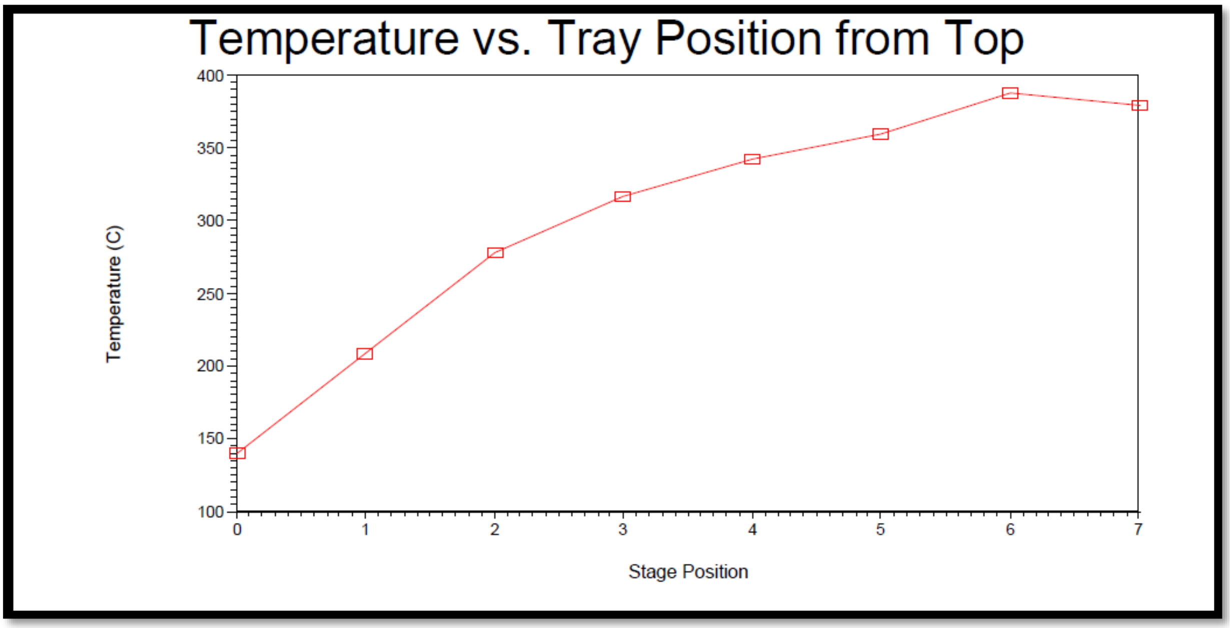

Figure 4.83: Temperature profile for blend4 (Mix1)

(Figure 4.84) showing the increasing of pressure during process.

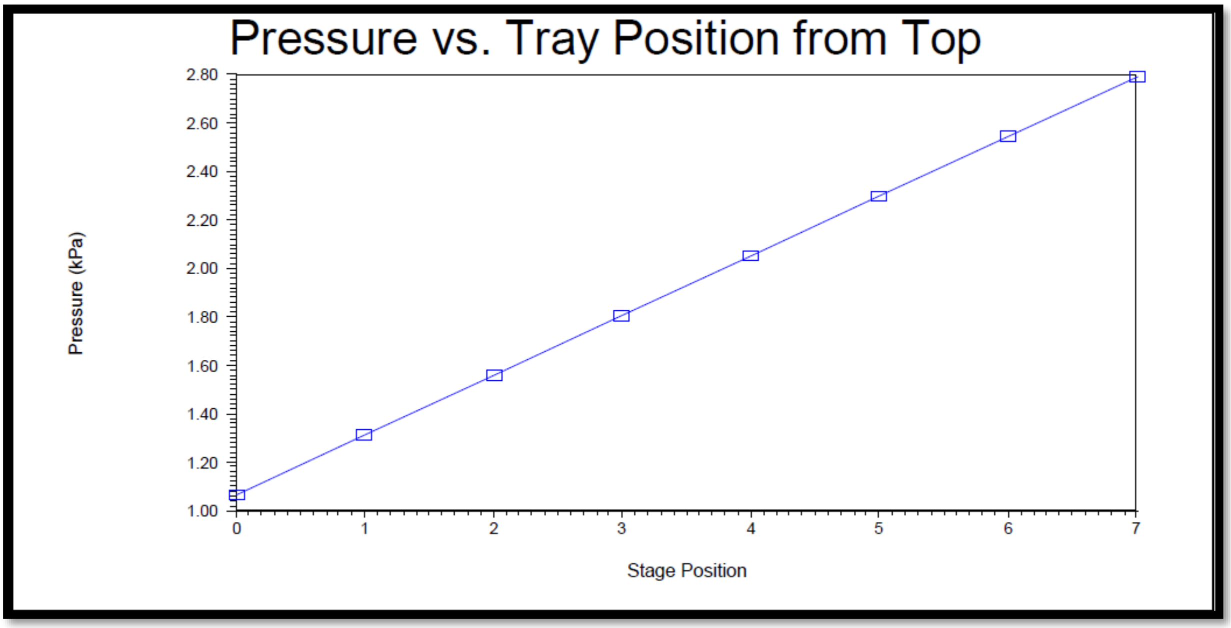

Figure 4.84: Pressure profile for blend4(Mix1)

The quantity of liquid flow and Vapour flow shown in (figure 4.85).

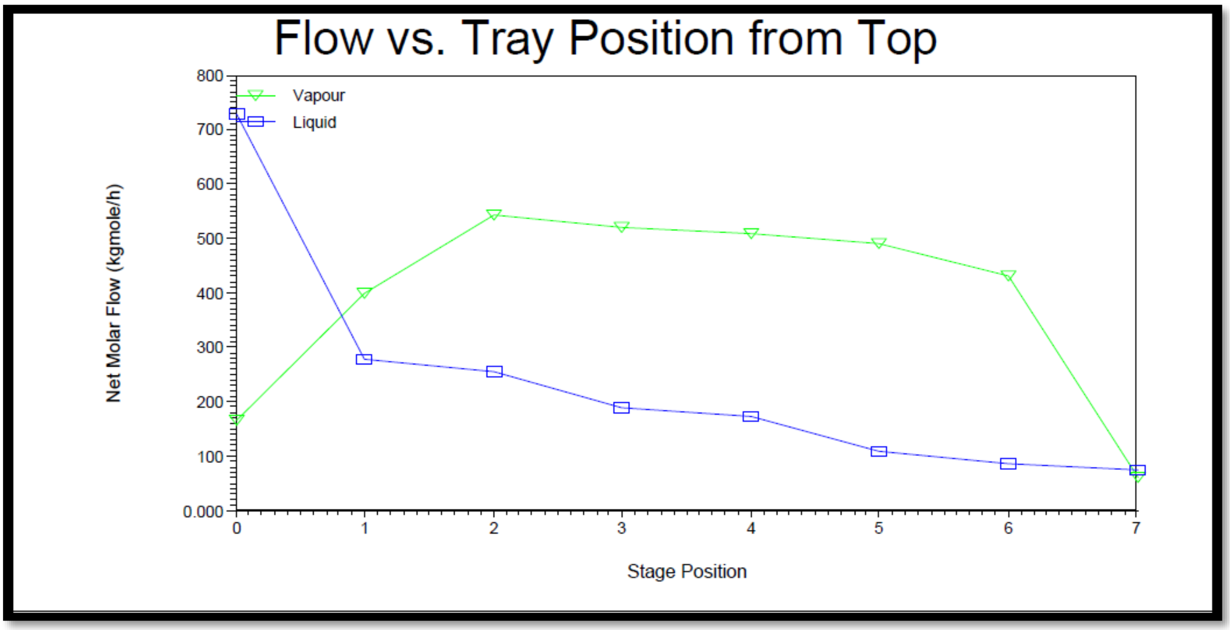

Figure 4.85: Net flow profile for blend4(Mix1)

column properties (Density and Molecular weight) shown in (figure 4.86).

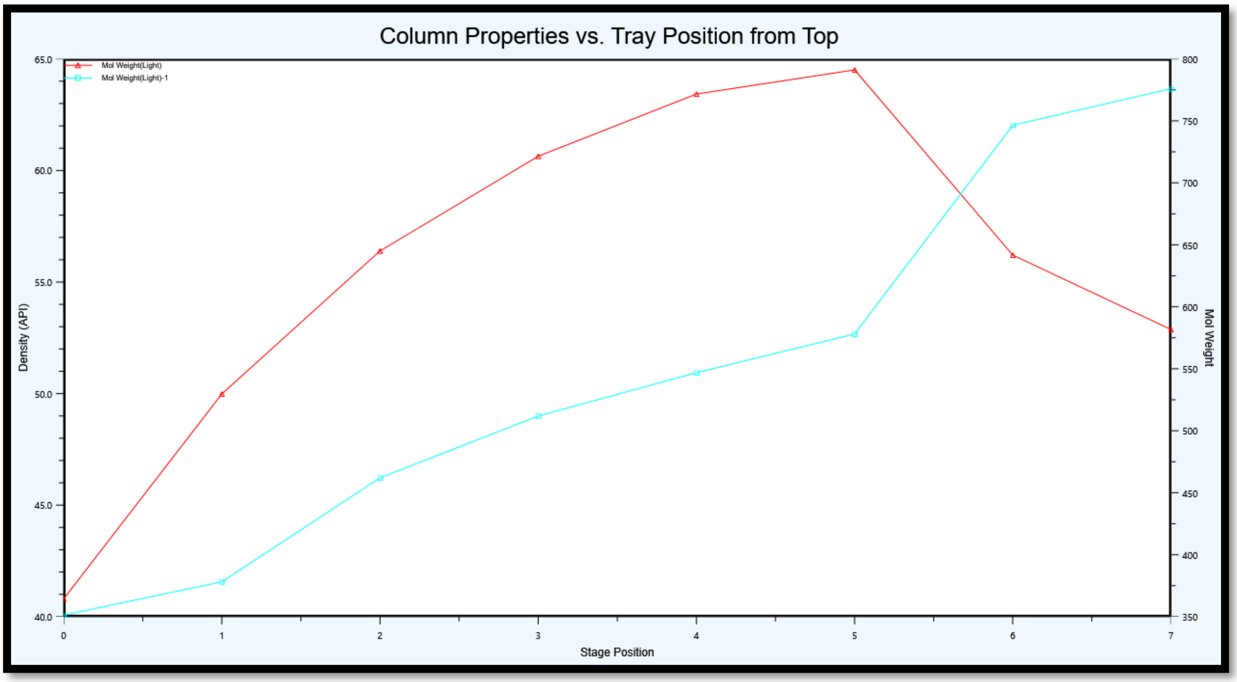

Figure 4.86: Correlation between density and molecular weight for blend4(Mix1)

4.6.3 Case Study no.3

Residue of the stream Mix 3 was chosen for entering in the vacuum distillation tower to convert it to light products.

The change of temperature from tray to another shown in (figure 4.87).

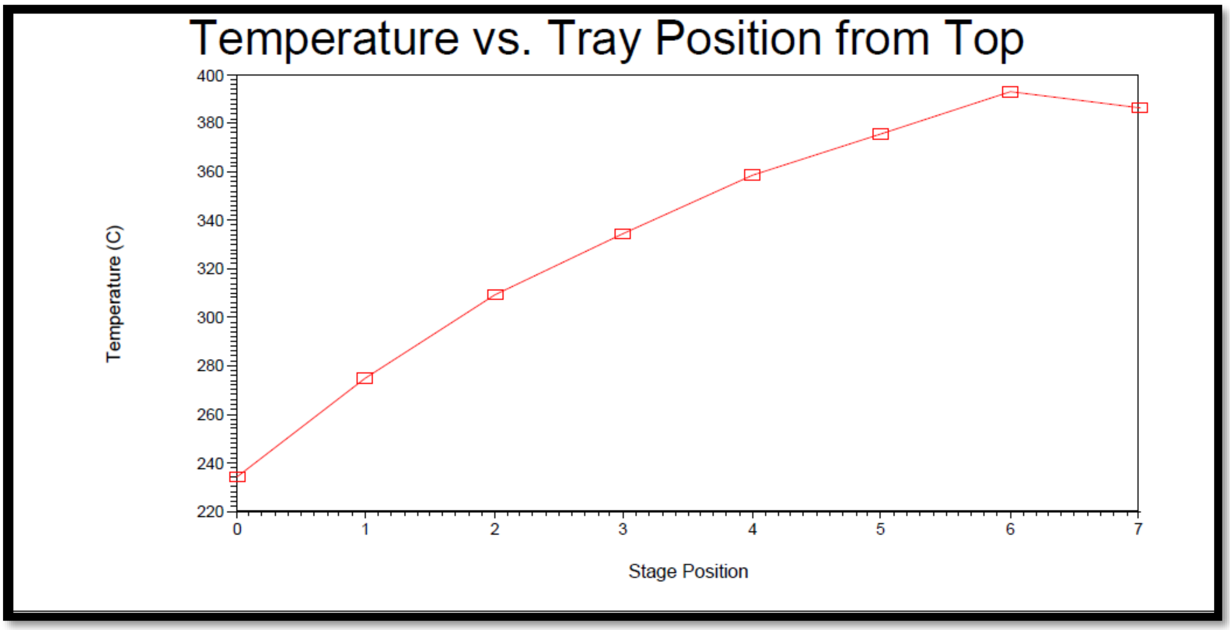

Figure 4.87: Temperature profile for blend6 (Mix3)

(Figure 4.88) showing the increasing of pressure during process.

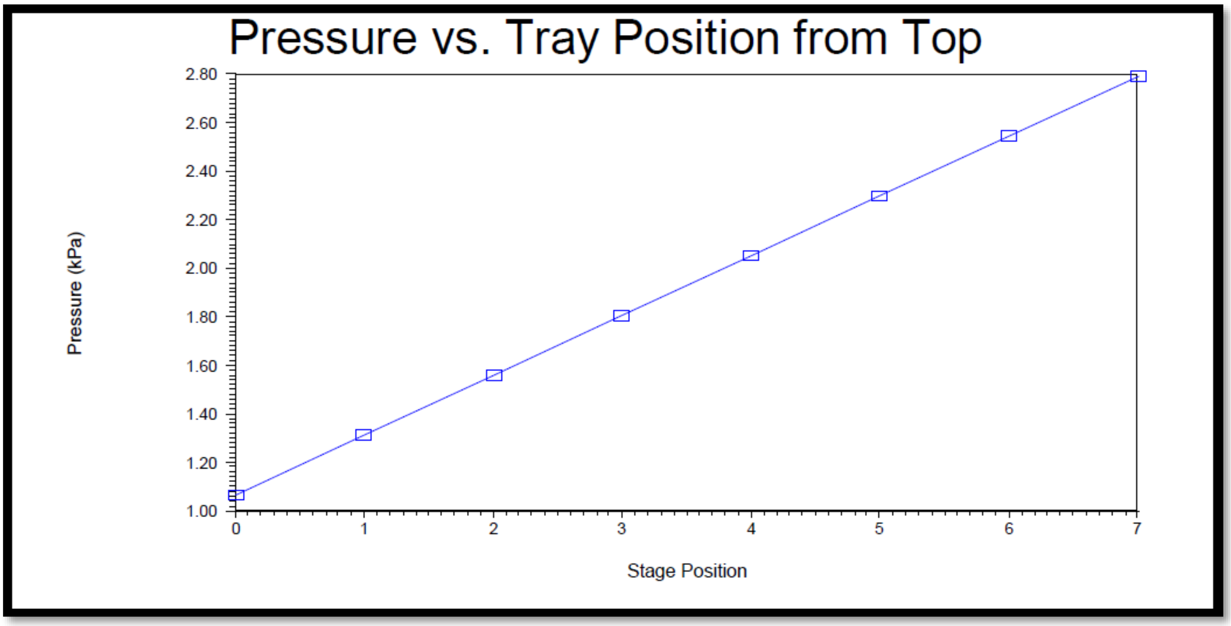

Figure 4.88: Pressure profile for blend6 (Mix3)

The quantity of liquid flow and Vapour flow shown in (figure 4.89).

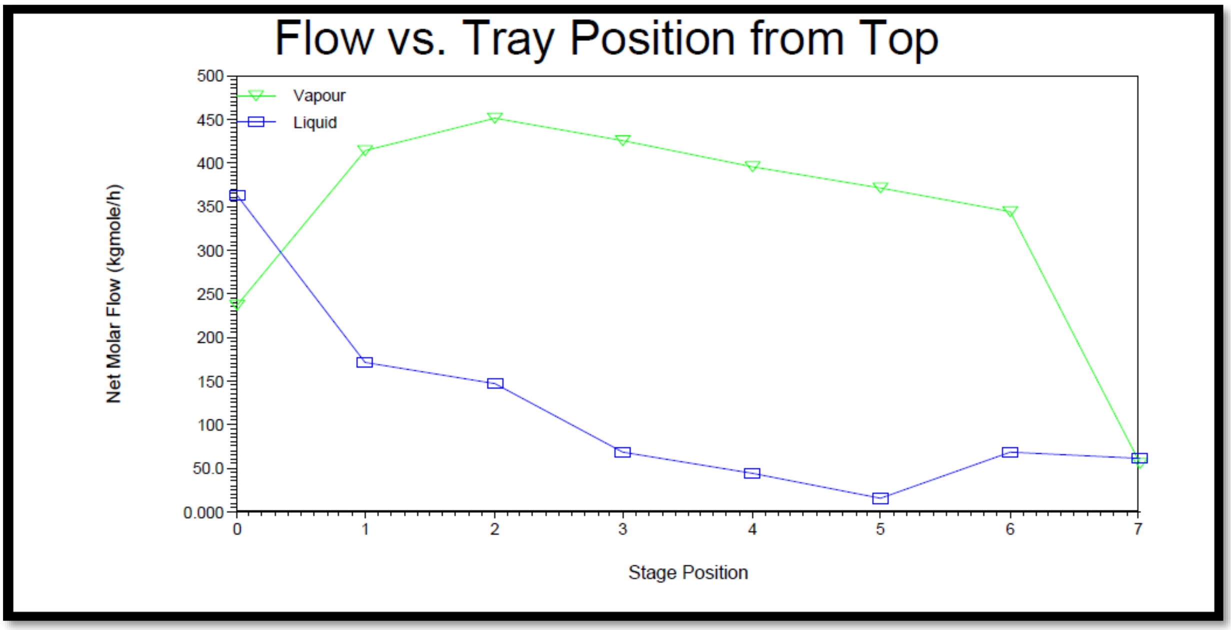

Figure 4.89: Net flow profile for blend6 (Mix3)

column properties (Density and Molecular weight) shown in (figure 4.90).

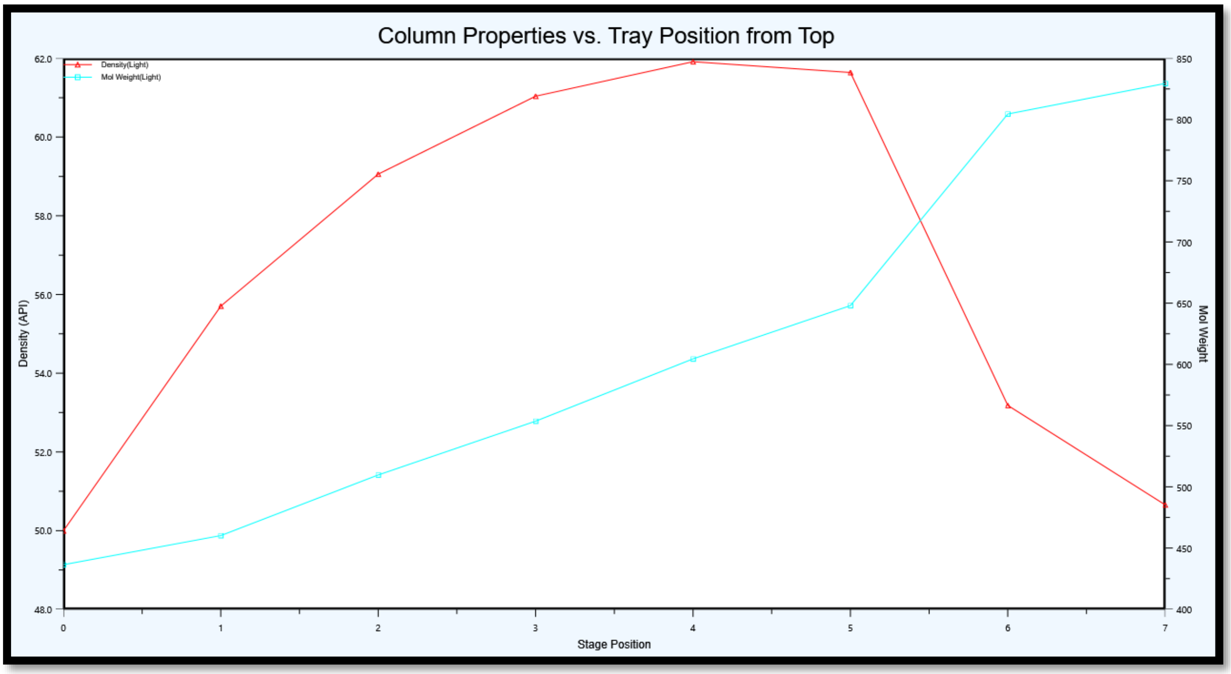

Figure 4.90: Correlation between density and molecular weight for blend6 (Mix3)

4.6.4 Case Study no.4

Residue of the stream Mix 7 was chosen for entering in the vacuum distillation tower to convert it to light products.

The change of temperature from tray to another shown in (figure 4.91).

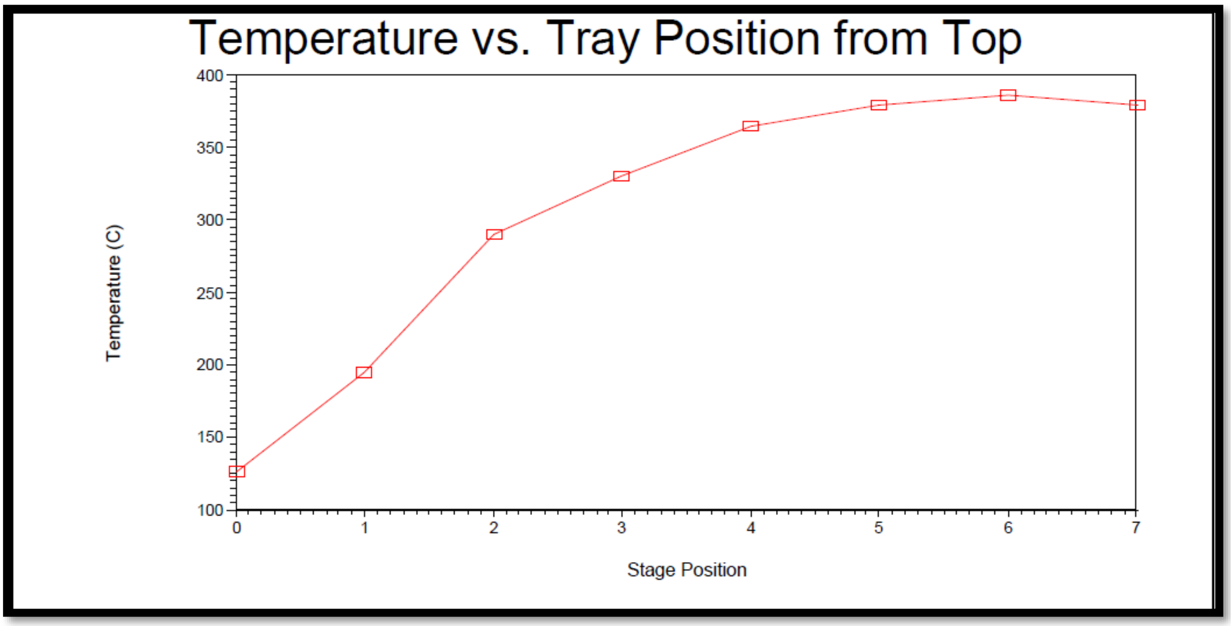

Figure 4.91: Temperature profile for blend10 (Mix7)

(Figure 4.92) showing the increasing of pressure during process.

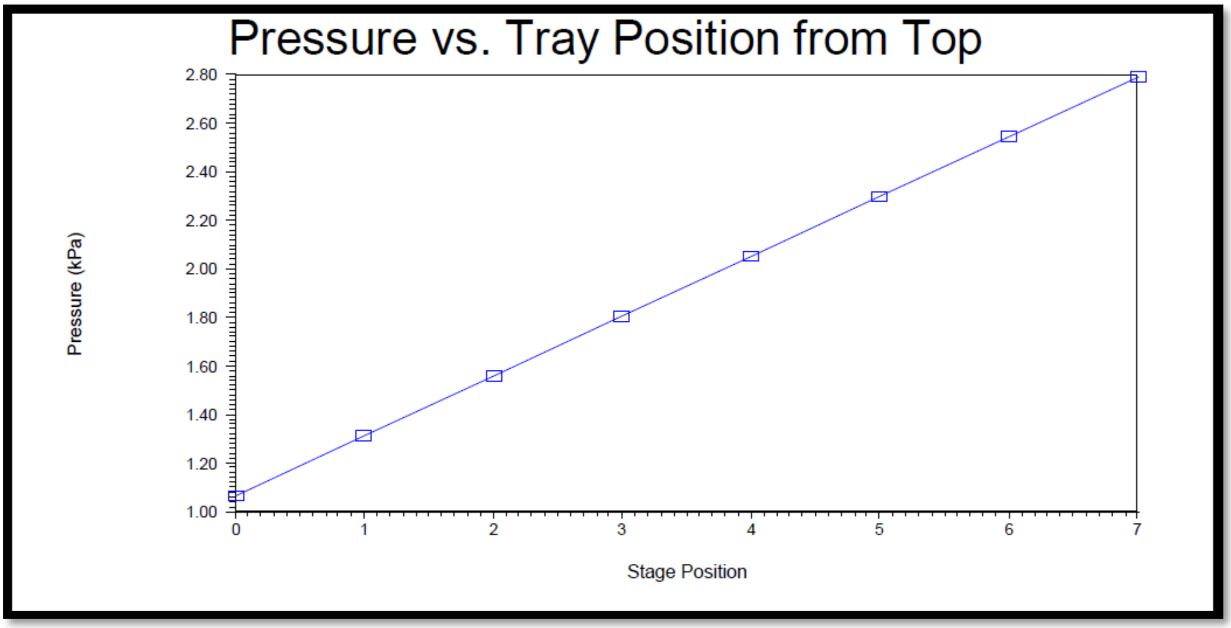

Figure 4.92: Pressure profile for blend10 (Mix7)

The quantity of liquid flow and Vapour flow shown in (figure 4.93).

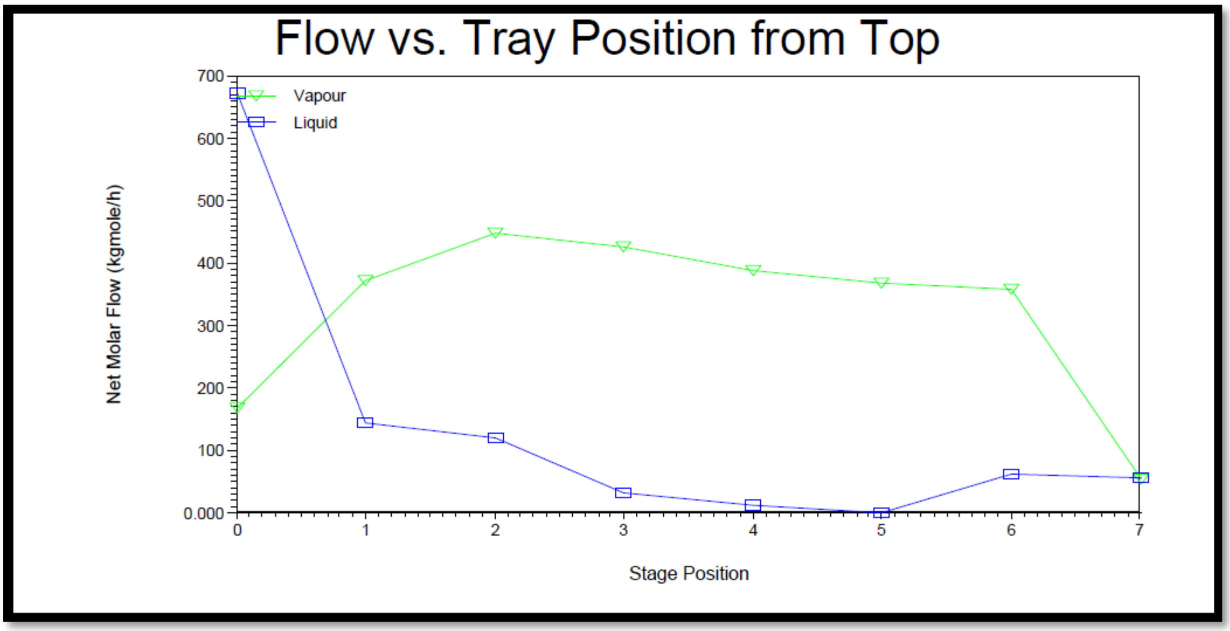

Figure 4.93: Net flow profile for blend10 (Mix7)

column properties (Density and Molecular weight) shown in (figure 4.94).

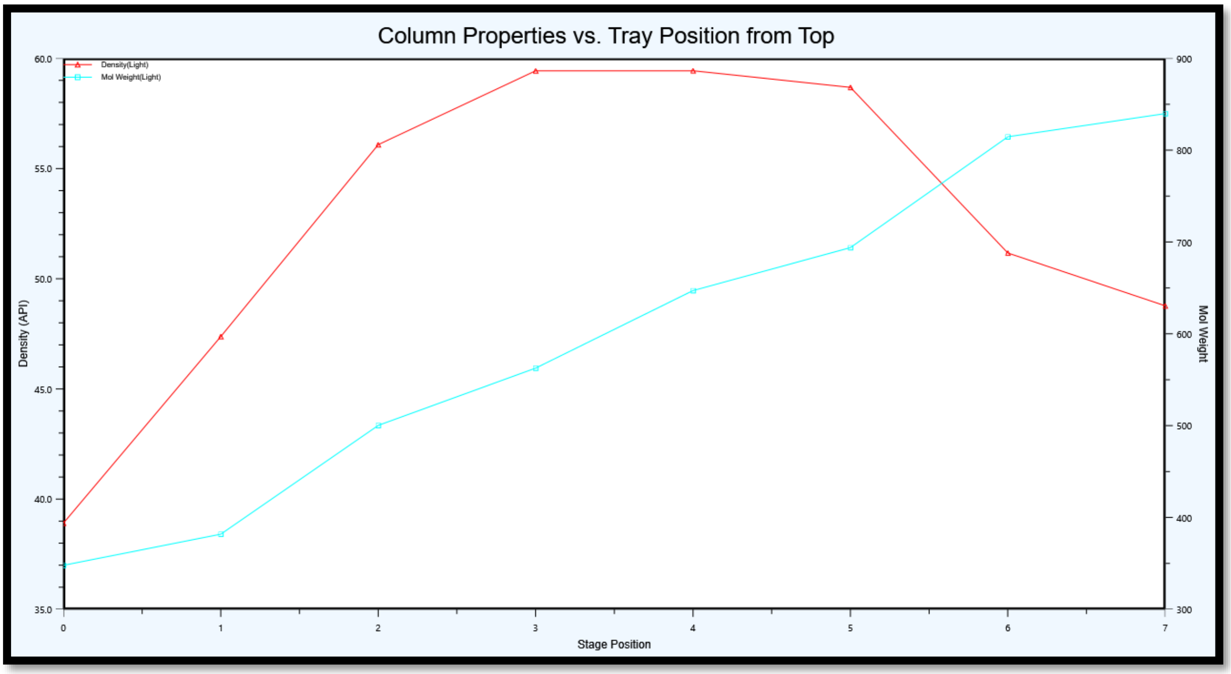

Figure 4.94: Correlation between density and molecular weight for blend10 (Mix7)

(Table 4.16) showing the flow rates products of vacuum distillation.

Table 4.16: The products of vacuum distillation unit

Type of crude Product	**Nile Blend** Blend 1	**Mix 1** Blend 4	**Mix 3** Blend 6	**Mix 7** Blend 10
Vacuum overhead rate	8.404 m^3/h	8.404 m^3/h	37.00 m^3/h	8.50 m^3/h
LVGO rate	33.93 m^3/h	41.36 m^3/h	21.03 m^3/h	57.93 m^3/h
HVGO rate	30.00 m^3/h	30.00 m^3/h	28.00 m^3/h	30.00 m^3/h
Slop wax rate	1.20 m^3/h	1.20 m^3/h	1.20 m^3/h	1.50 m^3/h
Vacuum Residue rate	66.34 m^3/h	59.90 m^3/h	52.19 m^3/h	48.36 m^3/h

Table 4.17: Boiling Point curve for streams

Properties **Streams**	**ASTM D93 Flash point**	**ASTM D97 Pour point**	**Viscosity at 37.8**	**Viscosity at 98.9**	**Paraffin Mole%**	**Naphthene Mole%**	**Aromatic Mole%**
Nile Blend	56.1985	20.9837	21.81905	5.96286	87.6267	12.3733	0.0000
Rawat	78.9083	20.4556	17.82911	3.91527	89.3168	10.6832	0.0000
Thargath	115.0213	33.6768	84.91345	12.85678	91.4796	8.5204	0.0000
Mix1	73.0513	23.8803	25.24903	6.20979	88.7938	11.2062	0.0000
Mix2	88.4740	26.7697	31.80275	7.09812	90.0541	9.9459	0.0000
Mix3	71.7855	23.8779	25.75626	6.36273	88.725	11.275	0.0000
Mix4	101.4060	29.4895	41.05403	8.19022	90.358	8.19022	0.0000
Mix5	80.6334	25.7943	30.01471	6.95686	89.6317	10.3683	0.0000
Mix6	69.8262	22.8816	23.48108	5.94893	88.402	11.598	0.0000
Mix7	65.0555	22.4031	23.3269	6.03579	88.107	11.893	0.0000

4.7 Comparison Between Atmospheric Residue and (VDU) products in boiling point curve (cold properties)

4.7.1 Case Study no.1 (Nile Blend)

(Table 4.18) showing the cold properties for product after Nile blend feed for atmospheric distillation tower, cold properties for product after Nile blend feed for vacuum distillation tower shown in (Table 4.19).

Table4.18: Boiling Point curve for product after Nile blend feed for atmospheric distillation tower

Product / **Properties**	Off Gas	Light Naphtha	Kerosene	Atmospheric Gas Oil	**Residue**
ASTM D93 Flash point	-51.8492	-26.4188	52.0195	98.3484	**151.0634**
ASTM D97 Pour point	0.0000	0.0000	-52.3475	-6.8791	**28.7590**
Cetane Index	0.0000	0.0000	31.9698	45.5165	**0.0000**
Research Octane Number	68.2911	61.6046	51.7764	0.0000	**0.0000**
Viscosity at 37.8	0.35688	0.40435	0.87321	3.62487	**251.86230**
Viscosity at 98.9	0.19591	0.23444	0.44882	1.13395	**20.50754**
Paraffin Mole%	72.0417	51.4743	37.6748	60.9529	**94.7179**
Naphthene Mole%	24.5476	33.3412	33.3762	27.1571	**5.2821**
Aromatic Mole%	3.4107	15.1845	28.9491	11.8899	**0.0000**

Table 4.19: Boiling Point curve for product after Nile blend feed for vacuum distillation tower

Product / Properties	Vacuum overhead	Light Vacuum Gas Oil	Heavy Vacuum Gas Oil	Slop Wax	Bottom Vacuum
ASTM D93 Flash point	106.4718	142.4752	0.0000	0.0000	0.0000
ASTM D97 Pour point	-8.2304	22.6332	31.4360	24.2076	-22.6044
Cetane Index	46.3557	0.0000	0.0000	0.0000	0.0000
Research Octane Number	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity at 37.8	1.72893	18.65811	166.09793	1181.10331	43996.99168
Viscosity at 98.9	0.72752	3.43725	11.69647	33.38874	415.46995
Paraffin Mole%	48.0570	78.7407	93.4258	96.9049	0.0000
Naphthene Mole%	34.2578	19.3080	6.5742	3.0951	0.0000
Aromatic Mole%	17.6852	1.9513	0.0000	0.0000	0.0000

4.7.2 Case Study no.2 (Mix 1)

(Table 4.20) showing the cold properties for product after Mix1 feed for atmospheric distillation tower, cold properties for product after Mix1 feed for vacuum distillation tower shown in (Table 4.21).

Table 4.20: Boiling Point curve for product after Mix1 feed for atmospheric distillation tower

Product **Properties**	Off Gas	Light Naphtha	Kerosene	Atmospheric Gas Oil	**Residue**
ASTM D93 Flash point	-51.8362	-17.8307	74.4506	112.2891	**152.1899**
ASTM D97 Pour point	0.0000	-87.3617	-35.1752	3.0260	**30.3150**
Cetane Index	0.0000	0.0000	39.2673	45.5273	**0.0000**
Research Octane Number	68.4070	60.5701	48.5968	0.0000	**0.0000**
Viscosity at 37.8	0.38076	0.49432	1.33869	5.51668	**233.09983**
Viscosity at 98.9	0.20909	0.28031	0.60852	1.48833	**18.45063**
Paraffin Mole%	73.1368	48.3372	36.5468	70.0486	**94.4391**
Naphthene Mole%	23.7422	33.2692	32.1920	22.2179	**5.5609**
Aromatic Mole%	3.1210	18.3936	31.2612	7.7335	**0.0000**

Table 4.21: Boiling Point curve for product after Mix1 feed for vacuum distillation tower

Product / Properties	Vacuum overhead	Light Vacuum Gas Oil	Heavy Vacuum Gas Oil	Slop Wax	Bottom Vacuum
ASTM D93 Flash point	111.9766	147.3093	0.0000	0.0000	0.0000
ASTM D97 Pour point	-2.4079	25.8492	30.6692	25.6864	-16.1529
Cetane Index	46.4032	0.0000	0.0000	0.0000	0.0000
Research Octane Number	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity at 37.8	1.90193	25.83612	248.64006	919.02191	32201.67395
Viscosity at 98.9	0.77753	4.29640	14.17386	28.71561	325.03392
Paraffin Mole%	63.5087	82.1713	94.2932	96.6133	0.0000
Naphthene Mole%	25.7750	17.0235	5.7068	3.3867	0.0000
Aromatic Mole%	10.7163	0.8052	0.0000	0.0000	0.0000

4.7.3 Case Study no.3 (Mix 3)

(Table 4.22) showing the cold properties for product after Mix3 feed for atmospheric distillation tower, cold properties for product after Mix3 feed for vacuum distillation tower shown in (Table 4.23).

Table4.22: Boiling Point curve for product after Mix 3 feed for atmospheric distillation tower

Product **Properties**	**Off Gas**	**Light Naphtha**	**Kerosene**	**Atmospheric Gas Oil**	**Residue**
ASTM D93 Flash point	-51.6044	-22.4412	66.8337	113.5521	**154.1144**
ASTM D97 Pour point	0.0000	-94.6800	-38.1475	3.9086	**29.1821**
Cetane Index	0.0000	0.0000	38.4419	45.2354	**0.0000**
Research Octane Number	68.0408	60.3412	50.5751	0.0000	**0.0000**
Viscosity at 37.8	0.36999	0.44289	1.24575	5.70594	**276.48301**
Viscosity at 98.9	0.20484	0.25536	0.57860	1.52905	**20.46315**
Paraffin Mole%	73.1258	50.9701	36.9091	69.9956	**94.8977**
Naphthene Mole%	23.7573	32.9133	32.2459	22.2434	**5.1023**
Aromatic Mole%	3.1169	16.1166	30.8449	7.7610	**0.0000**

Table 4.23: Boiling Point curve for product after Mix3 feed for vacuum distillation tower

Product / **Properties**	**Vacuum overhead**	**Light Vacuum Gas Oil**	**Heavy Vacuum Gas Oil**	**Slop Wax**	**Bottom Vacuum**
ASTM D93 Flash point	135.6022	162.8652	0.0000	0.0000	0.0000
ASTM D97 Pour point	20.2641	30.7423	27.6814	14.0493	-31.6702
Cetane Index	0.0000	0.0000	0.0000	0.0000	0.0000
Research Octane Number	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity at 37.8	10.12769	98.23853	573.33211	3551.49734	90810.85155
Viscosity at 98.9	2.15927	8.79668	22.14463	70.39902	717.53068
Paraffin Mole%	76.7434	92.8312	95.7747	0.0000	0.0000
Naphthene Mole%	20.6750	7.1688	4.2253	0.0000	0.0000
Aromatic Mole%	2.5816	0.0000	0.0000	0.0000	0.0000

4.7.4 Case Study no.4 (Mix 7)

(Table 4.24) showing the cold properties for product after Mix7 feed for atmospheric distillation tower, cold properties for product after Mix7 feed for vacuum distillation tower shown in (Table 4.25).

Table 4.24: Boiling Point curve for product after Mix 7 feed for atmospheric distillation tower

Product / properties	Off Gas	Light Naphtha	Heavy Naphtha	Kerosene	Atmospheric Gas Oil	Residue
ASTM D93 Flash point	-49.9937	-21.0306	38.3046	56.6254	102.6710	**147.0337**
ASTM D97 Pour point	0.0000	-96.4884	-64.7490	-47.1811	-4.8943	**30.5446**
Cetane Index	0.0000	0.0000	24.5034	33.5731	44.9687	**0.0000**
Research Octane Number	67.0030	59.4872	51.1474	51.4232	0.0000	**0.0000**
Viscosity at37.8	0.37784	0.42833	0.68358	0.99304	3.90573	**184.57298**
Viscosity at98.9	0.20870	0.24822	0.37124	0.49198	1.18586	**16.89104**
Paraffin Mole%	71.1350	51.2247	39.9300	37.4314	63.0772	**93.9736**
Naphthene Mole%	24.8416	33.0146	33.9774	32.8722	25.9459	**6.0264**
Aromatic Mole%	4.0234	15.7607	26.0926	29.6964	10.9769	**0.0000**

Table 4.25: Boiling Point curve for product after Mix7 feed for vacuum distillation tower

Product / **Properties**	**Vacuum overhead**	**Light Vacuum Gas Oil**	**Heavy Vacuum Gas Oil**	**Slop Wax**	**Bottom Vacuum**
ASTM D93 Flash point	103.1159	144.4691	0.0000	0.0000	0.0000
ASTM D97 Pour point	-10.1078	26.2648	26.2995	4.6330	-35.6738
Cetane Index	45.7250	0.0000	0.0000	0.0000	0.0000
Research Octane Number	0.0000	0.0000	0.0000	0.0000	0.0000
Viscosity at 37.8	1.69354	28.09333	684.02574	7652.54530	112328.76042
Viscosity at 98.9	0.71636	4.74747	24.81861	135.16282	835.25222
Paraffin Mole%	35.7664	83.2756	96.1949	0.0000	0.0000
Naphthene Mole%	31.3733	16.3304	3.8051	0.0000	0.0000
Aromatic Mole%	32.8603	0.3939	0.0000	0.0000	0.0000

4.8 Discussion:

In this study, the main problem was the urgent need for light products, this project aims to study and upgrading the heavy Sudanese crude oil and convert it into high-value light products through the Aspen Hysys program, and compare the results obtained with the basic crude (Nile Blend) used in El Obeid Refinery.

Firstly 7 blends of different heavy Sudanese crude oils were made, and its entered into a simulation program specially designed to simulate El Obeid refinery. The percentages of products (cut distribution) for all cases were just taken from TBP curve according to main distillation conditions and the results for the different streams were as follows:

Nile Blend: The conversion ratio to high-value light products in the Atmospheric distillation tower of the Nile Blend was 31%vol, and its fractions as follows:

0.7084 m^3/h off-gas (0.71%), 7.255 m^3/h Light Naphtha (7.3%),4.832 m^3/h Kerosene (4.86%), 18.03 m^3/h Atmospheric Gas Oil(AGO) (18.13%) and 68.62 m^3/h Residue (69%).

Rawat: The conversion ratio to high-value light products in the Atmospheric distillation tower of the Rawat were 39.7%vol, and its fractions as follows:

0.3287 m^3/h off-gas (1.65%), 3.00 m^3/h Light Naphtha (15.05%),0.585 m^3/h Kerosene (2.93%), 3.999 m^3/h Atmospheric Gas Oil(AGO) (20.06%) and 12.02 m^3/h Residue (60.3%).

Thargath: The conversion ratio to high-value light products in the Atmospheric distillation tower of the Thargath were 24.22%vol, and its fractions as follows: 0.00 m^3/h off-gas (0.00%), 1.80 m^3/h Light Naphtha (1.81%),4.869 m^3/h Kerosene (4.90%), 17.4 m^3/h Atmospheric Gas Oil(AGO) (17.51%) and 75.33 m^3/h Residue (75.78%).

Blend 4 (**Mix 1**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 1 were 30.51%vol, and its fractions as follows:

0.8007 m^3/h off-gas (0.80%), 7.867 m^3/h Light Naphtha (7.91%), 4.76 m^3/h Kerosene (4.79%), 16.92 m^3/h Atmospheric Gas Oil(AGO) (17.01%) and 69.12 m^3/h Residue (69.49%).

Blend 5 (**Mix 2**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 2 were 26.62%vol, and its fractions as follows:

0.5463 m^3/h off-gas (0.55%), 4.751 m^3/h Light Naphtha (4.78%), 5.872 m^3/h Kerosene (5.9%), 15.3 m^3/h Atmospheric Gas Oil(AGO) (15.39%) and 72.97 m^3/h Residue (73.38%).

Blend 6 (**Mix 3**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 3 were 31.23%vol, and its fractions as follows:

0.807 m^3/h off-gas (0.81%), 6.586 m^3/h Light Naphtha (6.62 %), 6.916 m^3/h Kerosene (6.95%), 16.75 m^3/h Atmospheric Gas Oil(AGO) (16.84%) and 68.40 m^3/h Residue (68.77%).

Blend 7 (**Mix 4**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 4 were 28.96%vol, and its fractions as follows:

0.9169 m^3/h off-gas (0.92%), 4.50 m^3/h Light Naphtha (4.52 %), 5.96 m^3/h Kerosene (6.00%), 17.44 m^3/h Atmospheric Gas Oil(AGO) (17.53%) and 70.68 m^3/h Residue (71.04%).

Blend 8 (**Mix 5**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 5 were 30.47%vol, and its fractions as follows:

0.7474 m^3/h off-gas (0.75%), 5.465 m^3/h Light Naphtha (5.49%), 7.043 m^3/h Kerosene (7.08%), 17.05 m^3/h Atmospheric Gas Oil(AGO) (17.14%) and 69.15 m^3/h Residue (69.53%).

Blend 9 (**Mix 6**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 6 were 27.97%vol, and its fractions as follows:

0.2471 m^3/h off-gas (0.25%), 6.47 m^3/h Light Naphtha (6.51%), 6.00 m^3/h Kerosene (6.04%), 15.09 m^3/h Atmospheric Gas Oil(AGO) (15.18%) and 71.61 m^3/h Residue (72.03%).

Blend 10 (**Mix 7**): The conversion ratio to high-value light products in the Atmospheric distillation tower of the Mix 7 were 27.81%vol, and its fractions as follows:

1.389 m^3/h off-gas (1.4%), 6.148 m^3/h Light Naphtha (6.18%), 5.136 m^3/h Kerosene (5.16%), 15.00 m^3/h Atmospheric Gas Oil(AGO) (15.07%) and 71.83 m^3/h Residue (72.19%).

Under normal distillation conditions, Rawat crude should produce more volume of off- gas while Thargath crude should give the least because it was heavy.

All comparisons are made with Nile Blend crude, the most converged crudes with the Nile Blend crude for produce off-gas are mix 1, mix 3, mix 5 and mix7(mix 7 produce more volume than Nile Blend of off-gas).

Rawat crude should produce more volume of Light Naphtha while Thargath crude should give the least because it was heavy, the most converged crudes with the Nile Blend crude for produce Light Naphtha are mix 1, mix 3, mix 6 and mix 7(mix 1 produce more volume than Nile Blend of Light Naphtha).

Mix 5 crude should produce more volume of Kerosene while Rawat crude should give the least, the most converged crudes with the Nile Blend crude for produce Kerosene are Thargath, mix 1, mix 3 and mix 7 (mix 3 produce more volume than Nile Blend of Kerosene).

Rawat crude should produce more volume of Atmospheric Gas Oil while Mix 7 crude should give the least, the most converged crudes with the Nile Blend crude for produce Atmospheric Gas Oil are Thargath, mix 1, mix 4 and mix5.

Thargath crude should produce more volume of Atmospheric Residue while Rawat crude should give the least, the most converged crudes with the Nile Blend crude for produce Atmospheric Residue are mix 1, mix 3 and mix 5.

The results obtained from the distribution of the oil were compared by volume percentage, and it was found that the results were most closely related to the Nile Blend were mix 1, mix 3 and mix 7. The conversion ratio in the atmospheric distillation tower to light products was 31.00% Nile Blend, 30.51% mix 1 ,31.23% mix 3 and 27.81% mix 7.

In the second stage, and before the design of the vacuum distillation unit, another atmospheric distillation column was added with the same specifications as the first tower to increase the amount of residual entering the atmospheric distillation tower to convert it into light products.

The characteristics of the different mixtures were recording before fed the atmospheric distillation tower shown in (Table 4.17) and record again after leaving the both atmospheric distillation tower(products) shown in tables (4.18,4.20,4.22 and 4.24), most of properties decrease (flash point, pour point, viscosity and paraffin) after leaving the both atmospheric distillation tower except residue and the Naphthene and aromatic increase except residue.

Vacuum distillation furnace may be classified into two types, wet and dry, the best method for vacuum process was wet type, steam was injected into the furnace coils and that helps to lower the partial pressure of feed as well as steam carries the feed vapours through the furnace tube more rapidly.

Proportional integral controller was selected to use for the system for Controlling of the temperature of vacuum furnace by control of the fuel flowrate, it provides a minimum overshoot than P and PID controllers in changing the step and making the system stable.

To improve vaporization, the effective pressure is lowered even further (to 27.58 kpa or less) by the addition of steam to the furnace inlet and at the bottom of the vacuum tower (Shukran and Abd Rahman.2013).

After that a comparison was made between the residue produced from the atmospheric distillation tower and the feed to the vacuum distillation tower and the products of the vacuum distillation tower in the properties and extraction ratios for each of the Nile Blend (Blend1), Mix 1(Blend4), Mix 3 (Blend6) and Mix 7 (Blend10).

When the residual of atmospheric distillation towers was fed to the vacuum distillation tower, its properties are compared with the products of the vacuum distillation tower as follows:

The characteristics of the flash point decrease in the light product (vacuum overhead, LVGO) and are not valued in Heavy products (slop wax, vacuum residue), paraffin and Naphthene decreasing in light products and increase in heavy products, while aromatic products appear only in the upper products of the vacuum distillation tower.

The conversion ratio to high-value light products in the vacuum distillation tower of the Nile Blend was 52.57%vol, and its fractions as follows:

6% vacuum overhead, 24.26% LVGO, 21.45% HVGO and 0.86% slop wax.

The conversion ratio for stream Mix1 to light products was 57.48% by volume, and its fractions as follows:

5.97% vacuum overhead, 29.36% LVGO, 21.3% HVGO and 0.85% slop wax.

The conversion ratio for stream Mix3 to light products was 62.57%by volume, and its fractions as follows:

26.54% vacuum overhead,15.08% LVGO, 20.08% HVGO and 0.86% slop wax.

The conversion ratio for stream Mix7 light products was 66.94% by volume, and its fractions as follows:

5.81% Vacuum overhead, 39.6% LVGO, 20.5% HVGO, and 1.03% Slop Wax.

CHAPTER FIVE

Conclusions and Recommendations

5.1 Conclusions

Crude oil is a term used to describe hydrocarbon rich mixtures that accumulated over millions of years and are usually found underground. This crude can be converted to other more useful products. The objective in refinery operations is to generate as much profit as possible by converting crude oils into valuable products such as gasoline, jet fuel, diesel, and so on.

The market need for these products is ever increasing all the time, which will necessarily cause a thinking to increase the production volume of the refining industry.

Simulation is presently a mature and well tested technology and it is widely used for a variety of purposes, including design, control, test, optimization, and integration of process plants.

Simulation software is one of the best tools for a crude oil refinery. This can be used during the conceptual design as well during the entire life span of the equipment's. Aspen Hysys program enables the simulation of very complex crude distillation systems to be easily manner. The goal is to achieve by using Aspen Hysys program, which provide capability to design the entire process accurately. This study aimed to make the maximum use of the available resources (Sudanese heavy crude oil) and turn them into desirable products to fill the existing shortage, and add some proposals for the development of El Obeid Refinery.

Seven mixtures were made, in addition to the three basic crude oils, and they became as follows: (Nile Blend, Rawat, Thargath, mix 1, mix 2, mix 3, mix 4, mix 5, mix6 and mix 7), they were entered into the Aspen Hysys program under the same conditions as the atmospheric distillation tower of El Obeid refinery and the residue ratio (bottom of the tower) for streams as follow: (69.06%, 60.49%, 75.81%, 69.56%, 73.43%, 68.83%, 71.13%, 69.59%, 72.06% and 72.29%) by volume respectively.

The best method for vacuum process was wet type, steam was injected into the furnace coils and that helps to lower the partial pressure of feed as well as steam carries the feed vapours through the furnace tube more rapidly.

To improve vaporization, the effective pressure is lowered even further (to 27.58 k pa or less) by the addition of steam to the furnace inlet and at the bottom of the vacuum tower (Shukran and Abd Rahman .2013).

Proportional integral controller was selected to use for the system, it provides a minimum overshoot than P and PID controllers in changing the step and making the system stable.

Four streams were selected from them to fed their residues to vacuum distillation tower, and the residues were converted to high-value light products. The selected streams were (Nile Blend, mix 1, mix 3 and mix 7) and the percentage of converting it into light products (52.57%, 57.48% ,62.57% and 66.94%) by volume respectively.

5.2 Recommendations

- An in-depth research studies and Investigation of the different heavy crude and residue processing technique to select the most appropriate technically and economically.
- New studies about Adding other units in Sudanese refineries (FCC & RFCC) to convert marginal heavy crude oil into light, higher value crude.
- Recommend the officials in El Obeid refinery can study this research to determine the suitability of such method to upgrade El Obeid refinery.

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APPENDIXES

APPENDIX A

Laboratory Report ID:16972
Assay of Nile blend

True Boiling Point Distillation Data

Cut Number	Yield %Weight	Cum. % Weight	Yield %Volume	Cum. % Volume
C4	0.06	0.06	0.09	0.09
C4 – 60	0.45	0.52	0.60	0.70
60 – 75	0.46	0.98	0.59	1.29
75 – 90	0.82	1.80	1.02	2.31
90 – 105	0.70	2.50	0.85	3.16
105 – 120	0.82	3.32	0.99	4.14
120 – 135	0.80	4.12	0.94	5.09
135 – 149	0.95	5.06	1.11	6.20
149 – 165	1.05	6.11	1.22	7.42
165 – 180	0.91	7.03	1.04	8.45
180 – 195	1.24	8.27	1.39	9.85
195 – 210	0.71	8.98	0.79	10.63
210 – 225	2.15	11.13	2.35	12.98
225 – 232	0.86	11.99	0.93	13.91
232 – 240	0.71	12.70	0.77	14.68
240 – 255	0.71	13.41	0.76	15.44
255 – 270	2.19	15.60	2.33	17.77
270 – 285	1.61	17.21	1.71	19.47
285 – 300	3.30	20.51	3.44	22.91
300 – 315	2.23	22.74	2.33	25.24
315 – 330	2.27	25.01	2.35	27.59
330 – 345	2.06	27.07	2.13	29.72
345 – 360	2.71	29.78	2.78	32.51
360 – 369	1.37	31.15	1.40	33.91
369 – 400	0.80	31.96	0.82	34.73
400 – 420	3.69	35.65	3.75	38.48
420 – 440	4.95	40.61	5.00	43.48
440 - 460	5.36	45.97	5.36	48.84
460 – 480	2.63	48.59	2.60	51.44
480 - 500	3.69	52.29	3.64	55.08

CRUDE ASSAY REBORT
Rawat Petroleum Co. Ltd
Rawat C4 DST1

True Boiling Point Distillation Data

Cut No	Cut Range(C)	YIELD ON CRUDE						Density@15C (kg/L)
		WEIGHT (g)	WEIGHT		Volume (mL)	VOLEME %		
			CUT	CUMUL		CUT	CUMUL	
1	C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0000
2	35-50	31.0	0.4	0.4	49.0	0.6	0.6	0.6347
3	50-70	34.0	0.5	0.9	50.0	0.6	1.3	0.6753
4	70-90	50.0	0.7	1.7	70.0	0.9	2.2	0.7178
5	90-120	34.0	0.5	2.1	46.0	0.6	2.7	0.7336
6	120-140	59.0	0.8	3.0	79.0	1.0	3.7	0.7454
7	140-155	46.0	0.7	3.6	61.0	0.8	4.5	0.7569
8	155-170	64.0	0.9	4.6	83.0	1.1	5.6	0.7687
9	170-185	45.0	0.6	5.2	58.0	0.7	6.3	0.7768
10	185-200	54.0	0.8	6.0	69.0	0.9	7.2	0.7841
11	200-220	85.0	1.2	7.2	107.0	1.4	8.6	0.7901
12	220-240	108.0	1.6	8.8	135.0	1.7	10.3	0.7974
13	240-260	137.0	2.0	10.7	169.0	2.2	12.4	0.8056
14	260-280	147.0	2.1	12.8	180.0	2.3	14.7	0.8133
15	280-300	180.0	2.6	15.4	220.0	2.8	17.5	0.8186
16	300-320	185.0	2.7	18.1	225.0	2.9	20.4	0.8244
17	320-330	90.0	1.3	19.4	109.0	1.4	21.8	0.8290
18	330-350	180.0	2.6	22.0	216.0	2.8	24.6	0.8344
19	350-370	205.0	2.9	24.9	244.0	3.1	27.7	0.8394
20	370-435	775.0	11.1	36.0	921.5	11.7	39.4	0.8444
21	435-450	370.2	5.3	41.4	436.8	5.6	45.0	0.8494
22	450-470	396.8	5.7	47.1	466.1	5.9	50.9	0.8535
23	470-500	463.4	6.7	53.7	543.3	6.9	57.8	0.8568
24	500-565	484.7	7.0	60.7	559.3	7.1	65.0	0.8650
25	565+	2737.8	39.3	100	2747.6	35.0	100.0	0.9899

Laboratory Report No 13-008376
Greater Nile Petroleum Operating Co. Ltd
Thargath Assay Report

True Boiling Point Distillation Data

Cut Range C	Volume Percent		Weight Percent		Density@ 15 C Kg/L	K Factor
	Initial Yield	Final Yield	Initial Yield	Final Yield		
70-85	0.19	0.30	0.15	0.24	0.7418	11.56
85-100	0.30	0.46	0.24	0.36	0.7476	11.63
100-115	0.46	0.65	0.36	0.51	0.7511	11.73
115-135	0.65	0.92	0.51	0.74	0.7563	11.81
135-150	0.92	1.16	0.74	0.93	0.7643	11.86
150-170	1.16	1.64	0.93	1.33	0.7786	11.80
170-185	1.64	2.08	1.33	1.71	0.7956	11.69
185-205	2.08	2.74	1.71	2.29	0.8123	11.59
205-225	2.74	3.66	2.29	3.10	0.8269	11.55
225-245	3.66	5.21	3.10	4.50	0.8405	11.50
245-265	5.21	7.33	4.50	6.44	0.8545	11.45
265-285	7.33	9.79	6.44	8.71	0.8597	11.52
285-305	9.79	12.59	8.71	11.31	0.8655	11.58
305-325	12.59	15.62	11.31	14.17	0.8778	11.54
325-345	15.62	18.74	14.17	17.12	0.8816	11.61
345-365	18.74	21.82	17.12	20.06	0.8914	11.61
365-385	21.82	25.00	20.06	23.12	0.8965	11.66
385-400	25.00	27.50	23.12	25.54	0.9015	11.69
400-415	27.50	30.12	25.54	28.09	0.9053	11.73
415-430	30.12	32.82	28.09	30.72	0.9077	11.78
430-445	32.82	35.53	30.72	33.36	0.9098	11.83
445-460	35.53	38.23	33.36	36.01	0.9118	11.89
460-475	38.23	40.94	36.01	38.66	0.9138	11.94
475-490	40.94	43.56	38.66	41.23	0.9154	11.99
490-505	43.56	45.99	41.23	43.62	0.9160	12.06
505-520	45.99	48.45	43.62	46.05	0.9161	12.14
520-535	48.45	50.92	46.05	48.47	0.9162	12.21
535-550	50.92	53.37	48.47	50.89	0.9165	12.28
550-565	53.37	55.82	50.89	53.29	0.9170	12.34
565+	55.82	100.00	53.29	100.00	0.9852	12.35

APPENDIX B

C 103 DATA
(KEROSINE STRIPPER)

Tray no	1-4	
Tray diameter	60 cm	
Number of down comers per tray	1.0	
Down comer area	L	32.8 cm
	H	17.0 cm
Down comer weir height	7.0 cm	
Down comer clearance	6.0 cm	
Number of valves per tray	14	
Number of open valves	14	
Number of closed valves	0.0	
Valve open area	16.25 cm^2	
Opening area per tray	14x16.25 = 227.5 cm^2	
Distance between trays	61 cm	

El Obeid Refinery
Technology Section
Crude Fractionating Tower
Main Column-C-101

Description: -

1-C.101 outside data	
1	Crude fractionating tower =7 1.D.* 91 S.S w/ 2:1S.E heads
2	Diameter I.D. =2.12 m
3	Height =27.6 m
4	Material =SA-516-70 and 316 S.S
5	Number of trays =36
6	Trays type =Valve type
7	Code =ASME sec. VIII, Div.1. latest edition
8	Design pressure =75 psig + 1/8"C.A.
9	Design Temp. =750 ° F
10	Man ways =2-20 " w/ deviated cover
11	Nozzles =As required 150 # and 300# ANSI RF
12	Insulation =2" w/aluminum jacket
13	Top thickness =9.1 - 10.01 mm
14	Bottom thickness =17.0- 18.4 mm
15	CR pipe inlet diameter =10"
16	CR inlet Location =Tray No.4
17	Overhead line diameter =12"
18	L.Naphath reflux pipe diameter =4"
19	L.Naphath reflux Location =Tray No.36
20	H.N draw off line diameter =6"
21	H.N reflux pipe diameter =4"
22	H.N Vapor line diameter =3"
23	H.N Draw off Location =Tray No. 32

24	H.N reflux location = Tray No.34 and 33	
25	H.N vapor location =Tray No.35 and 34	
26	Kero. Draw off line diameter =4"	
27	Kero. Vapor line diameter =3"	
28	Kero. Draw off location =Tray No.22	
29	Kero. Vapor location =Tray no.27	
30	G/O draw off line diameter =6"	
31	G/O reflux pipe diameter =4"	
32	G/O vapor line diameter =3"	
33	G/O draw off location =Tray No. 13	
34	G/O reflux location =Tray No. 18 and 17	
35	G/O vapor location =Tray No. 19	
36	L. Residue outlet pipe diameters =6"	
37	Stripping steam pipe inlet diameters =2"	
38	Stripping steam inlet location =Tray No. 2 and 1	
2- C-101 in side data		
1-Tray No: 1-15		
1	Tray Diameter = 210.0 cm	
2	Number of Down comers per tray = 1.0	
3	Down comer area	L = 140.0cm
		H = 25.3 cm
4	Down comer weir height = 7.0 cm	
5	Down comer clearance = 6.5 cm	
6	Number of valve per tray = 219.0	
7	Number of open valve = 155.0	
8	Number of closed valve = 64.0	
9	Valve open area = 31.2 cm^2	
10	Opening area per tray =155 x 31.2 = 4836 cm^2	
11	Distance between trays = 76.5 cm	

2-Tray No: 16-20		
1	Tray Diameter = 210.0 cm	
2	Number of Down comers per tray = 1.0	
3	Down comer area	L = 120.4 cm
		H = 18.1 cm
4	Down comer weir height = 7.3 cm	
5	Down comer clearance = 5.8 cm	
6	Number of valve per tray = 191.0	
7	Number of open valve = 143.0	
8	Number of closed valve = 48.0	
9	Valve open area = 31.2 cm^2	
10	Opening area per tray =143x 31.2 = 4461.6 cm^2	
11	Distance between trays = 76.5 cm	
3-Tray no: 21-36		
1	Tray Diameter = 210.0 cm	
2	N of Down comers per tray = 1.0	
3	Down comer area	L = 120.4 cm
		H = 18.0 cm
4	Down comer weir height = 7.2 cm	
5	Down comer clearance = 5.6 cm	
6	Number of valve per tray = 191.0	
7	Number of open valve = 100.0	
8	Number of closed valve = 90.0	
9	Valve open area = 31.2 cm^2	
10	Opening area per tray =100 x 31.2 = 3120.0 cm^2	
11	Distance between tray 21-24 = 76.0 cm	
12	Distance between tray 24-32 = 60.0 cm	
13	Distance between tray 32-33 = 104.0 cm (flashing area)	
14	Distance between tray 33-36 = 46.0 cm	

El Obeid Refinery
Technology Section
Heaters

DATA:

(1) Heaters type =Horizontal direct fired
(2) Heaters area =(L.W.H) (10,4,5.5) m
(3) Heater Insulation = Old heaters
(4) Flow arrangement =Single pass
(5) Av.flux rate = 10,000 Btu/hr. /Ft^2
(6) Design capacity = 1350 T/day
(7) Operating capacity = 1400 T/day
(8) Max capacity = 2020 T/day
(9) Heater duty = 16.0MMBtu/hr
(10) Stack height = 10.6m

(11) Tubes: -

1- Tubes type = Horizontal
2- length = 9.55m
3- Diameter = 4"
4- Material = C.S.SA-106-Bsmls.pipe
5- Elbow (U.tubes elb 180°) =4"
6- Number of tubes = 56
7- Tubes: 40 Radiant -4"Sch.40A-106-B. Bare
8- Tubes: 8 convection-4"Sch-40A-106-B Finned
9- Tubes: 4 convection-4"Sch-40A-106-B Finned
10-Tubes: 4 steam preheat-4"Sch-40A-106-B finned

11-CR &steam

12- CR inlet temp- =420°F
13- CR outlet temp- =700°F
14- Operating CR inlet pressure =125 psig
15- CR pipe inlet diameter = 4"
16- CR pipe outlet diameter =4"
17- Steam pipe inlet and outlet diameter =4"
18- Steam decoking line =2"
19- Steam inlet to burner =2” ___ 1"
20- Fuel inlet to burner =2” ___ ½"
21- Butane gas inlet to burner =1” _____ ½"
22- Steam to steaming out line = 2"
23- Number of thermocouples per heater =8

24- Heater temp- and pressure by thermocouple: -

1- Heaters inlet (CR) =420°F
2- Heaters outlet (CR) =700°F
3- Lower left tube skin =954°F
4- Lower right tube skin =1030°F

5- Upper left tube skin =959°F
6- Upper right tube skin =914°F
7- Left fire box =1430°F
8- Right fire box = 1428°F
9- East stack =1025°F
10- West stack =973°F
11- Steam to heaters temp. =356°F
12- Steam from heaters temp. (S.H.steam) =744°F
13- Steam pressure to heaters =82 psig

25- Fuel and steam consumption: -

1- Type of fuel =Sudanese crude oil
2- Fuel consumption =21.924 m ton/day
3- Steam to heaters consumption = 1.46 ton/day
4- Butane gas to heaters = 16.4kg/day

26- Heat transfer method: -

1-Convection
2-Radiation

27- **Heaters Duty: -**

1- Mass flow =128625Ib/hr
2- Specific heat (C.P) =0.73 Btu/hr
3- Temp.outlet/inlet =700,420°F
4-Q=128625(700-420) 0.73 =26290950 Btu/hr
5- Heaters duty is to heat 128625 Ib/hr of Crude oil
From 420°F to 700°F (26290050 Btu/hr)

28- **Burner Data: -**

1- Diameter =2.0mm
2- Number of nozzle =6
3- Gun length = 0.3m
4- Model & size = Model MA-16
5- Number of burner required = Eight (8)
6- Fuels fired =*6 oil
7- Tile included = Yes
8- Pilot = Quoted manual Ignition as option
9- Wind box detail =N/A
10- Noise attenuation =N/A
11- Surface prep./finish =SSPC.SP2 with one (1) coat alkyd primer
12- Tile required =HEVD-T-V-9
13- Pilot operating pressures =2.5. psig

14- **Heat release per burner (net): -**

Max = 7,500,000 Btu/hr
Normal = 5,000,000 Btu/hr
Min = 2,000,000 Btu/hr

15- Pilot orifice bridled = 1/16" Ø
16- Fuel gas = 910Btu/ScF (LHV)
17- Specific Gravity = 0.6
18- Dimension from heater steel to gas conn =24 ⅝"
19- Dimension from heater steel to pilot conn = 23¼"

Remark

* 6 oil viscosity must be 250 SSU. steam differential of 30 psig required.

29- Heater Data: -

1- Item No =oil fired burners
2- Type heater =Cabin
3- Fire box dimensions = 10,0 width x29, 11.25
4- Direction of firing = Horizontal (End wall)
5- Type of draft =natural
6- Setting thickness = Customer to advise

30- Operating Data: -

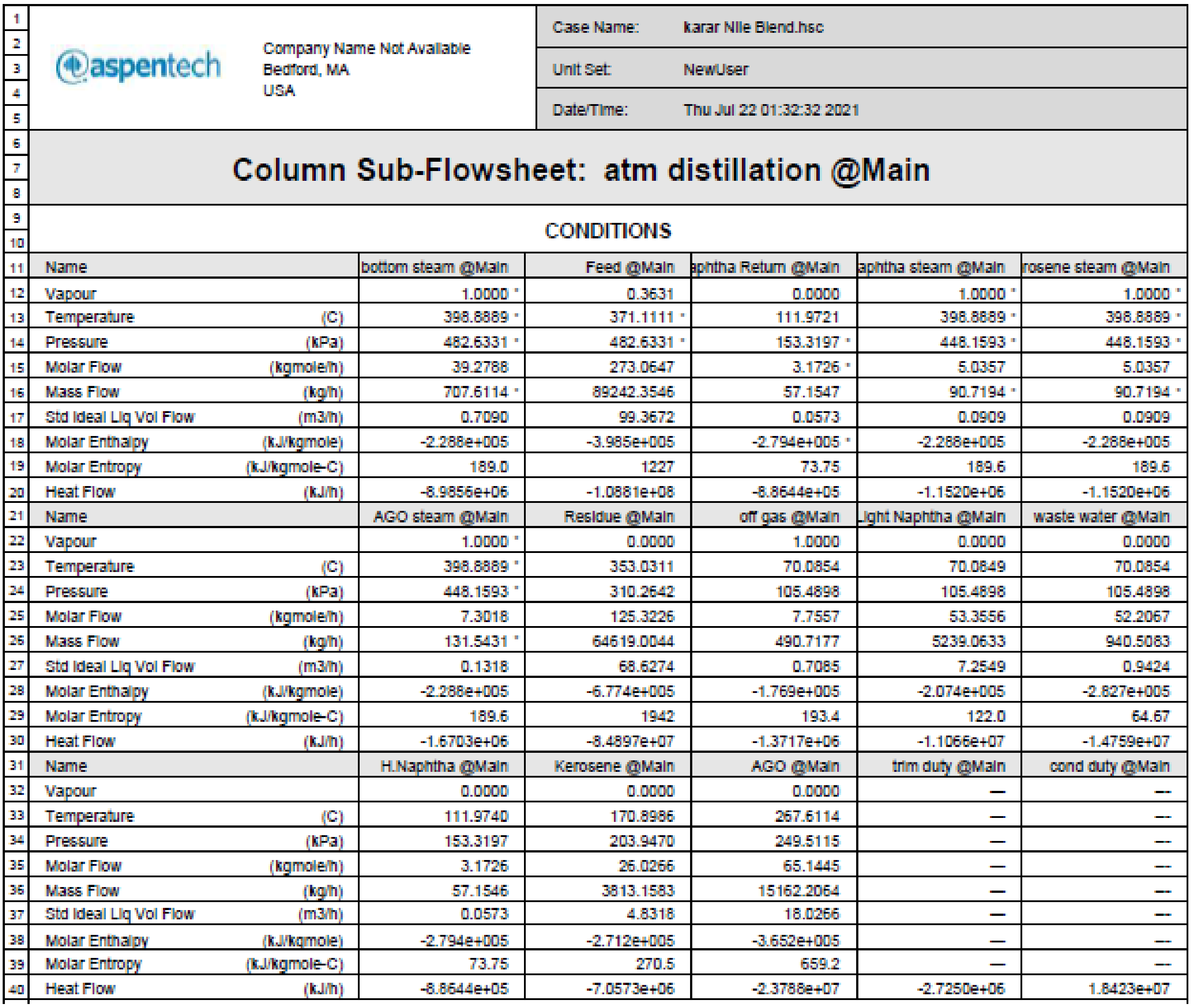
1-Heat release per burner: -
*Maximum =7.00 MM Btu/hr
*Normal = 5.00 MM Btu/hr
*Minimum = 2.33 MM Btu/hr
2- Turn down = 3:1
3- Excess air at maximum =15%

4-Conditions at burner: -

1- Pressure drop at burner = 0.30” W.C
2- Air temperature = 100°F
3- Fuel gas pressure available = 100 psig required
4- Estimated flame dimension =2.5" x 1.3"

APPENDIX C

Atmospheric Distillation: Nile Blend Conditions



Company Name Not Available
Bedford, MA
USA

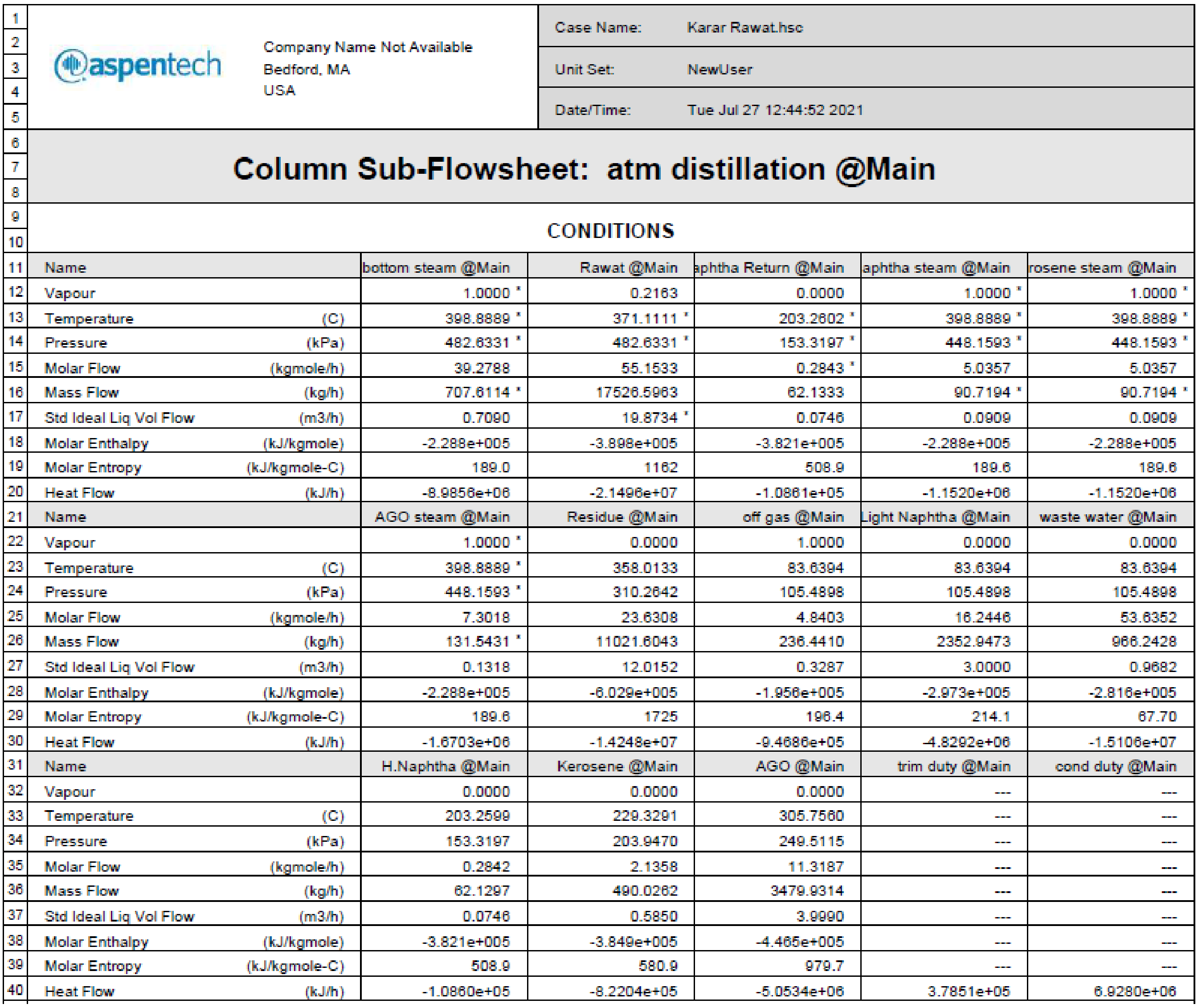
Case Name: karar Nile Blend.hsc
Unit Set: NewUser
Date/Time: Thu Jul 22 01:32:32 2021

Column Sub-Flowsheet: atm distillation @Main

CONDITIONS

Name		bottom steam @Main	Feed @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.3631	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	111.9721	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	273.0647	3.1726 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	89242.3546	57.1547	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672	0.0573	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-3.985e+005	-2.794e+005 *	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1227	73.75	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.0881e+08	-8.8644e+05	-1.1520e+06	-1.1520e+06
Name		AGO steam @Main	Residue @Main	off gas @Main	ight Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	353.0311	70.0854	70.0849	70.0854
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	125.3226	7.7557	53.3556	52.2067
Mass Flow	(kg/h)	131.5431 *	64619.0044	490.7177	5239.0633	940.5083
Std Ideal Liq Vol Flow	(m3/h)	0.1318	68.6274	0.7085	7.2549	0.9424
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.774e+005	-1.769e+005	-2.074e+005	-2.827e+005
Molar Entropy	(kJ/kgmole-C)	189.6	1942	193.4	122.0	64.67
Heat Flow	(kJ/h)	-1.6703e+06	-8.4897e+07	-1.3717e+06	-1.1066e+07	-1.4759e+07
Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	—	—
Temperature	(C)	111.9740	170.8986	267.6114	—	—
Pressure	(kPa)	153.3197	203.9470	249.5115	—	—
Molar Flow	(kgmole/h)	3.1726	26.0266	65.1445	—	—
Mass Flow	(kg/h)	57.1546	3813.1583	15162.2064	—	—
Std Ideal Liq Vol Flow	(m3/h)	0.0573	4.8318	18.0266	—	—
Molar Enthalpy	(kJ/kgmole)	-2.794e+005	-2.712e+005	-3.652e+005	—	—
Molar Entropy	(kJ/kgmole-C)	73.75	270.5	659.2	—	—
Heat Flow	(kJ/h)	-8.8644e+05	-7.0573e+06	-2.3788e+07	-2.7250e+06	1.8423e+07

Atmospheric Distillation: Rawat Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: Karar Rawat.hsc

Unit Set: NewUser

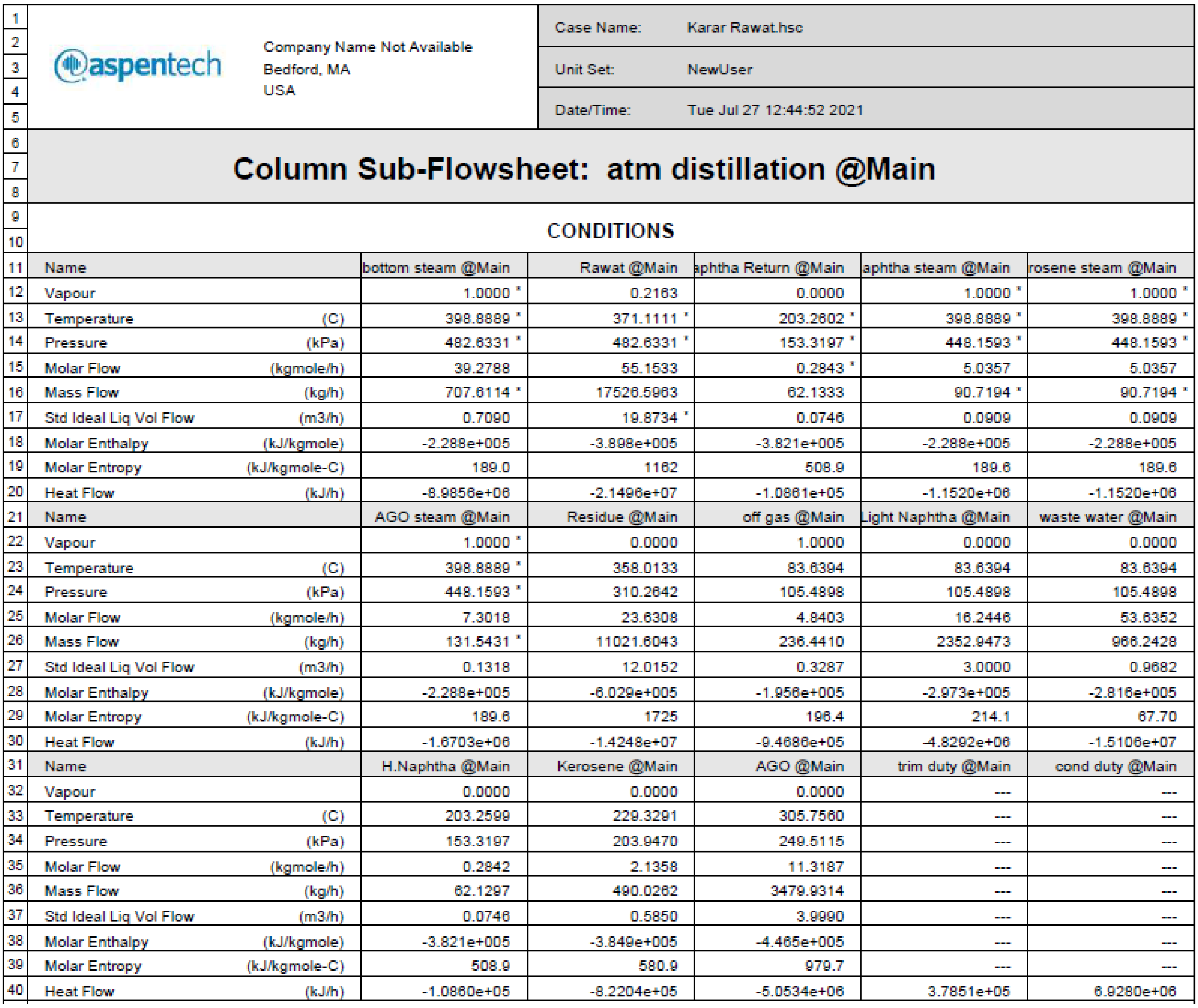
Date/Time: Tue Jul 27 12:44:52 2021

Column Sub-Flowsheet: atm distillation @Main

CONDITIONS

Name		bottom steam @Main	Rawat @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2163	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	203.2602 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	55.1533	0.2843 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	17526.5963	62.1333	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	19.8734 *	0.0746	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-3.898e+005	-3.821e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1162	508.9	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-2.1496e+07	-1.0861e+05	-1.1520e+06	-1.1520e+06
Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	358.0133	83.6394	83.6394	83.6394
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	23.6308	4.8403	16.2446	53.6352
Mass Flow	(kg/h)	131.5431 *	11021.6043	236.4410	2352.9473	966.2428
Std Ideal Liq Vol Flow	(m3/h)	0.1318	12.0152	0.3287	3.0000	0.9682
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.029e+005	-1.956e+005	-2.973e+005	-2.816e+005
Molar Entropy	(kJ/kgmole-C)	189.6	1725	196.4	214.1	67.70
Heat Flow	(kJ/h)	-1.6703e+06	-1.4248e+07	-9.4686e+05	-4.8292e+06	-1.5106e+07
Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	203.2599	229.3291	305.7560	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	0.2842	2.1358	11.3187	---	---
Mass Flow	(kg/h)	62.1297	490.0262	3479.9314	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0746	0.5850	3.9990	---	---
Molar Enthalpy	(kJ/kgmole)	-3.821e+005	-3.849e+005	-4.465e+005	---	---
Molar Entropy	(kJ/kgmole-C)	508.9	580.9	979.7	---	---
Heat Flow	(kJ/h)	-1.0860e+05	-8.2204e+05	-5.0534e+06	3.7851e+05	6.9280e+06

Atmospheric Distillation: Thargath Conditions



aspentech

Company Name Not Available
Bedford, MA
USA

Case Name:	Karar Rawat.hsc
Unit Set:	NewUser
Date/Time:	Tue Jul 27 12:44:52 2021

Column Sub-Flowsheet: atm distillation @Main

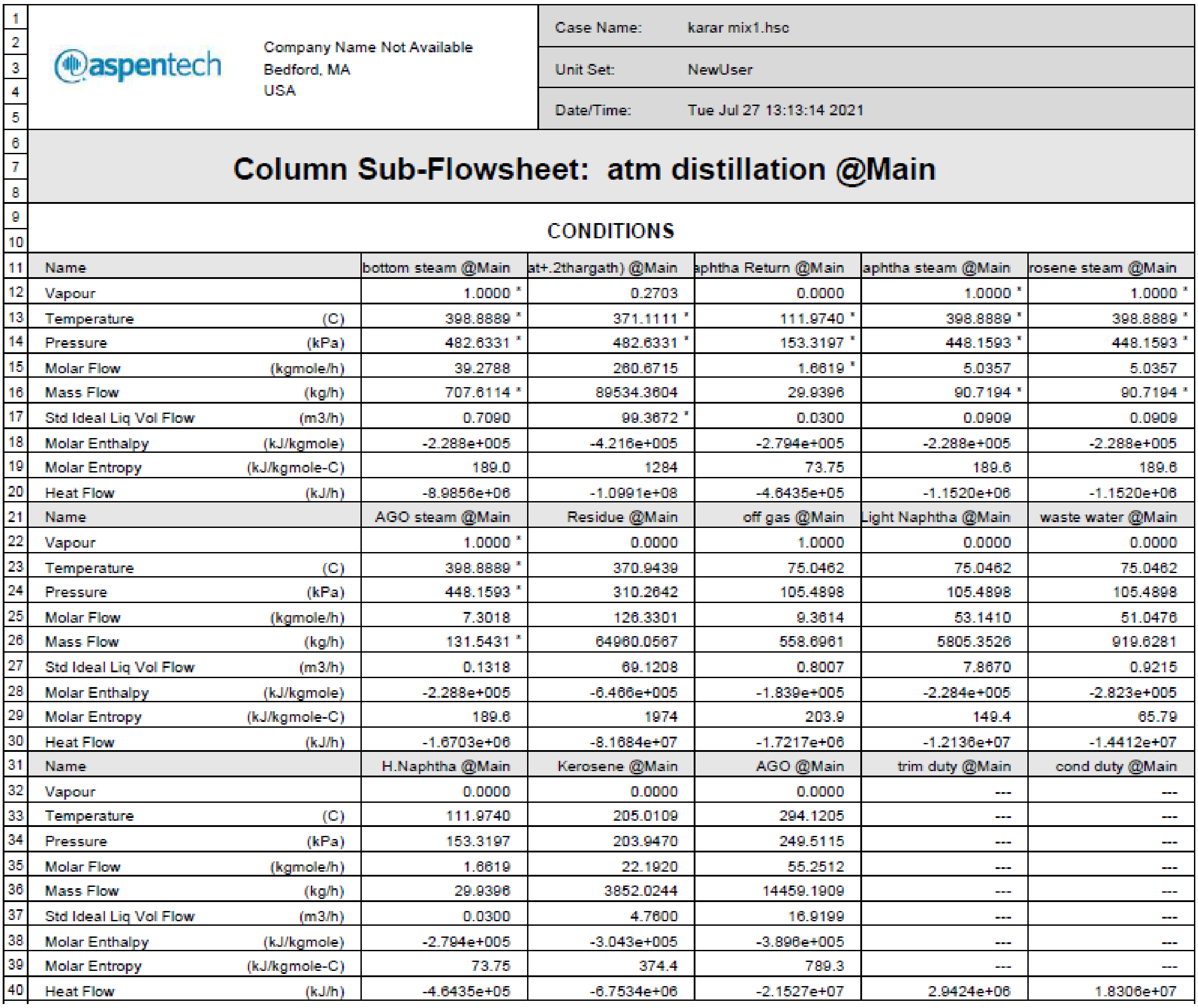
CONDITIONS

Name		bottom steam @Main	Rawat @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2163	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	203.2602 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	55.1533	0.2843 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	17526.5963	62.1333	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	19.8734 *	0.0746	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-3.898e+005	-3.821e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1162	508.9	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-2.1496e+07	-1.0861e+05	-1.1520e+06	-1.1520e+06

Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	358.0133	83.6394	83.6394	83.6394
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	23.6308	4.8403	16.2446	53.6352
Mass Flow	(kg/h)	131.5431 *	11021.6043	236.4410	2352.9473	966.2428
Std Ideal Liq Vol Flow	(m3/h)	0.1318	12.0152	0.3287	3.0000	0.9682
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.029e+005	-1.956e+005	-2.973e+005	-2.816e+005
Molar Entropy	(kJ/kgmole-C)	189.6	1725	196.4	214.1	67.70
Heat Flow	(kJ/h)	-1.6703e+06	-1.4248e+07	-9.4686e+05	-4.8292e+06	-1.5106e+07

Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	203.2599	229.3291	305.7560	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	0.2842	2.1358	11.3187	---	---
Mass Flow	(kg/h)	62.1297	490.0262	3479.9314	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0746	0.5850	3.9990	---	---
Molar Enthalpy	(kJ/kgmole)	-3.821e+005	-3.849e+005	-4.465e+005	---	---
Molar Entropy	(kJ/kgmole-C)	508.9	580.9	979.7	---	---
Heat Flow	(kJ/h)	-1.0860e+05	-8.2204e+05	-5.0534e+06	3.7851e+05	6.9280e+06

Atmospheric Distillation: Mix 1 Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: karar mix1.hsc

Unit Set: NewUser

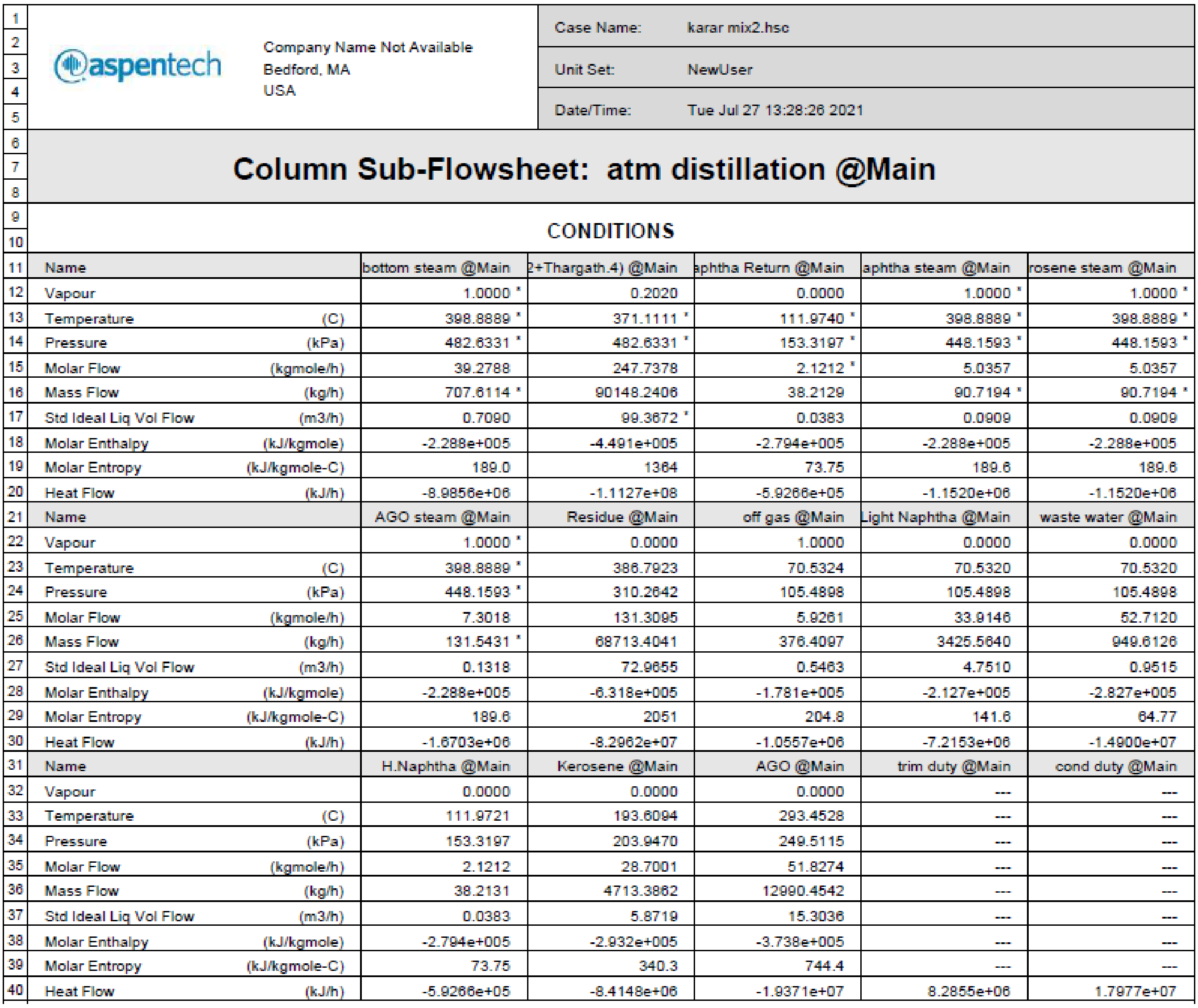
Date/Time: Tue Jul 27 13:13:14 2021

Column Sub-Flowsheet: atm distillation @Main

CONDITIONS

Name		bottom steam @Main	at+.2thargath) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2703	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	111.9740 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	260.6715	1.6619 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	89534.3604	29.9396	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.0300	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.216e+005	-2.794e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1284	73.75	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.0991e+08	-4.6435e+05	-1.1520e+06	-1.1520e+06
Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	370.9439	75.0462	75.0462	75.0462
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	126.3301	9.3614	53.1410	51.0476
Mass Flow	(kg/h)	131.5431 *	64960.0567	558.6961	5805.3526	919.6281
Std Ideal Liq Vol Flow	(m3/h)	0.1318	69.1208	0.8007	7.8670	0.9215
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.466e+005	-1.839e+005	-2.284e+005	-2.823e+005
Molar Entropy	(kJ/kgmole-C)	189.6	1974	203.9	149.4	65.79
Heat Flow	(kJ/h)	-1.6703e+06	-8.1684e+07	-1.7217e+06	-1.2136e+07	-1.4412e+07
Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	111.9740	205.0109	294.1205	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	1.6619	22.1920	55.2512	---	---
Mass Flow	(kg/h)	29.9396	3852.0244	14459.1909	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0300	4.7600	16.9199	---	---
Molar Enthalpy	(kJ/kgmole)	-2.794e+005	-3.043e+005	-3.896e+005	---	---
Molar Entropy	(kJ/kgmole-C)	73.75	374.4	789.3	---	---
Heat Flow	(kJ/h)	-4.6435e+05	-6.7534e+06	-2.1527e+07	2.9424e+06	1.8306e+07

Atmospheric Distillation: Mix 2 Conditions



aspentech

Company Name Not Available
Bedford, MA
USA

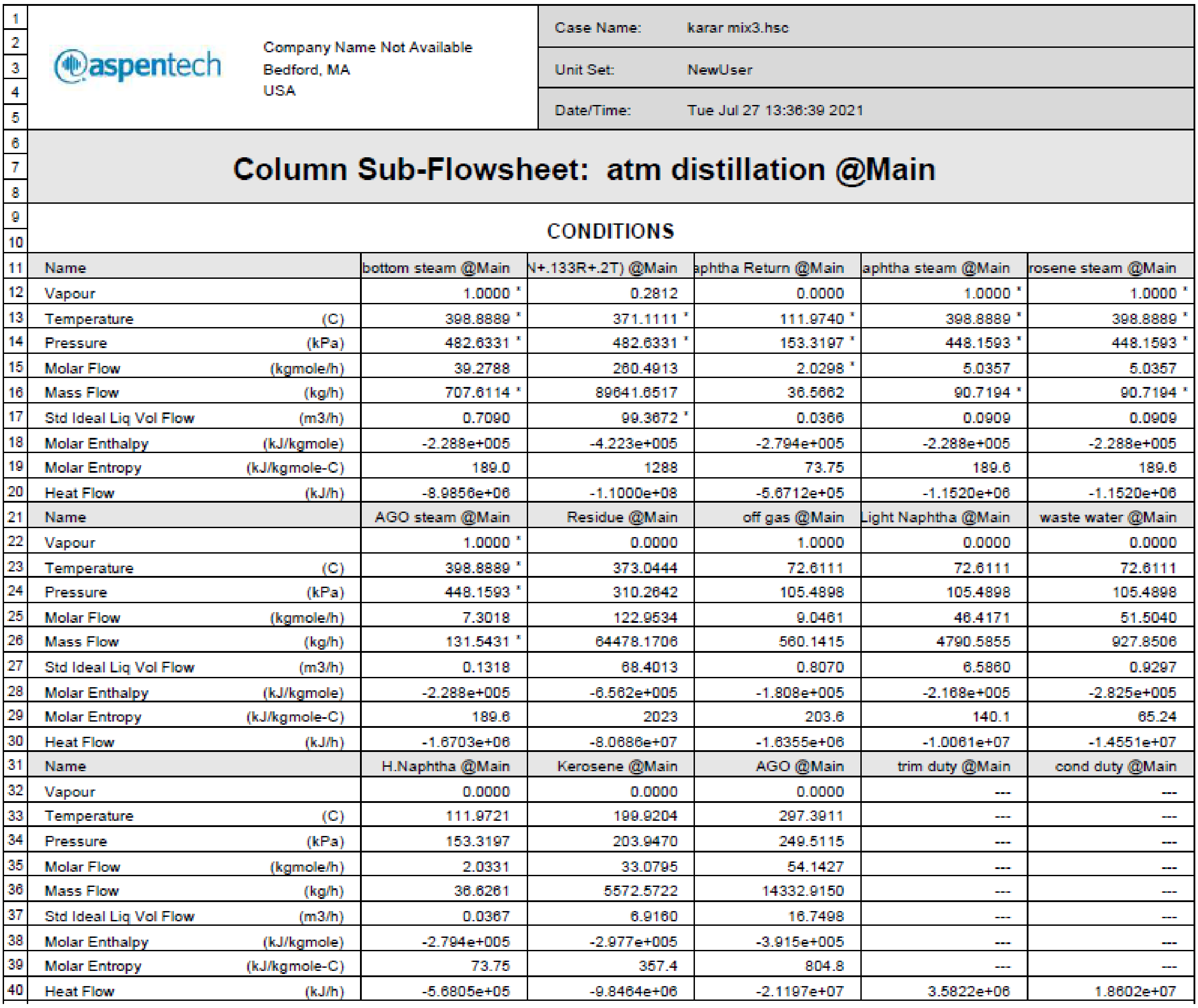
Case Name: karar mix2.hsc
Unit Set: NewUser
Date/Time: Tue Jul 27 13:28:26 2021

Column Sub-Flowsheet: atm distillation @Main

CONDITIONS

Name		bottom steam @Main	2+Thargath.4) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2020	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	111.9740 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	247.7378	2.1212 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	90148.2406	38.2129	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.0383	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.491e+005	-2.794e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1364	73.75	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.1127e+08	-5.9266e+05	-1.1520e+06	-1.1520e+06
Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	386.7923	70.5324	70.5320	70.5320
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	131.3095	5.9261	33.9146	52.7120
Mass Flow	(kg/h)	131.5431 *	68713.4041	376.4097	3425.5640	949.6126
Std Ideal Liq Vol Flow	(m3/h)	0.1318	72.9655	0.5463	4.7510	0.9515
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.318e+005	-1.781e+005	-2.127e+005	-2.827e+005
Molar Entropy	(kJ/kgmole-C)	189.6	2051	204.8	141.6	64.77
Heat Flow	(kJ/h)	-1.6703e+06	-8.2962e+07	-1.0557e+06	-7.2153e+06	-1.4900e+07
Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	111.9721	193.6094	293.4528	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	2.1212	28.7001	51.8274	---	---
Mass Flow	(kg/h)	38.2131	4713.3862	12990.4542	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0383	5.8719	15.3036	---	---
Molar Enthalpy	(kJ/kgmole)	-2.794e+005	-2.932e+005	-3.738e+005	---	---
Molar Entropy	(kJ/kgmole-C)	73.75	340.3	744.4	---	---
Heat Flow	(kJ/h)	-5.9266e+05	-8.4148e+06	-1.9371e+07	8.2855e+06	1.7977e+07

Atmospheric Distillation: Mix 3 Conditions



aspentech

Company Name Not Available
Bedford, MA
USA

Case Name: karar mix3.hsc
Unit Set: NewUser
Date/Time: Tue Jul 27 13:36:39 2021

Column Sub-Flowsheet: atm distillation @Main

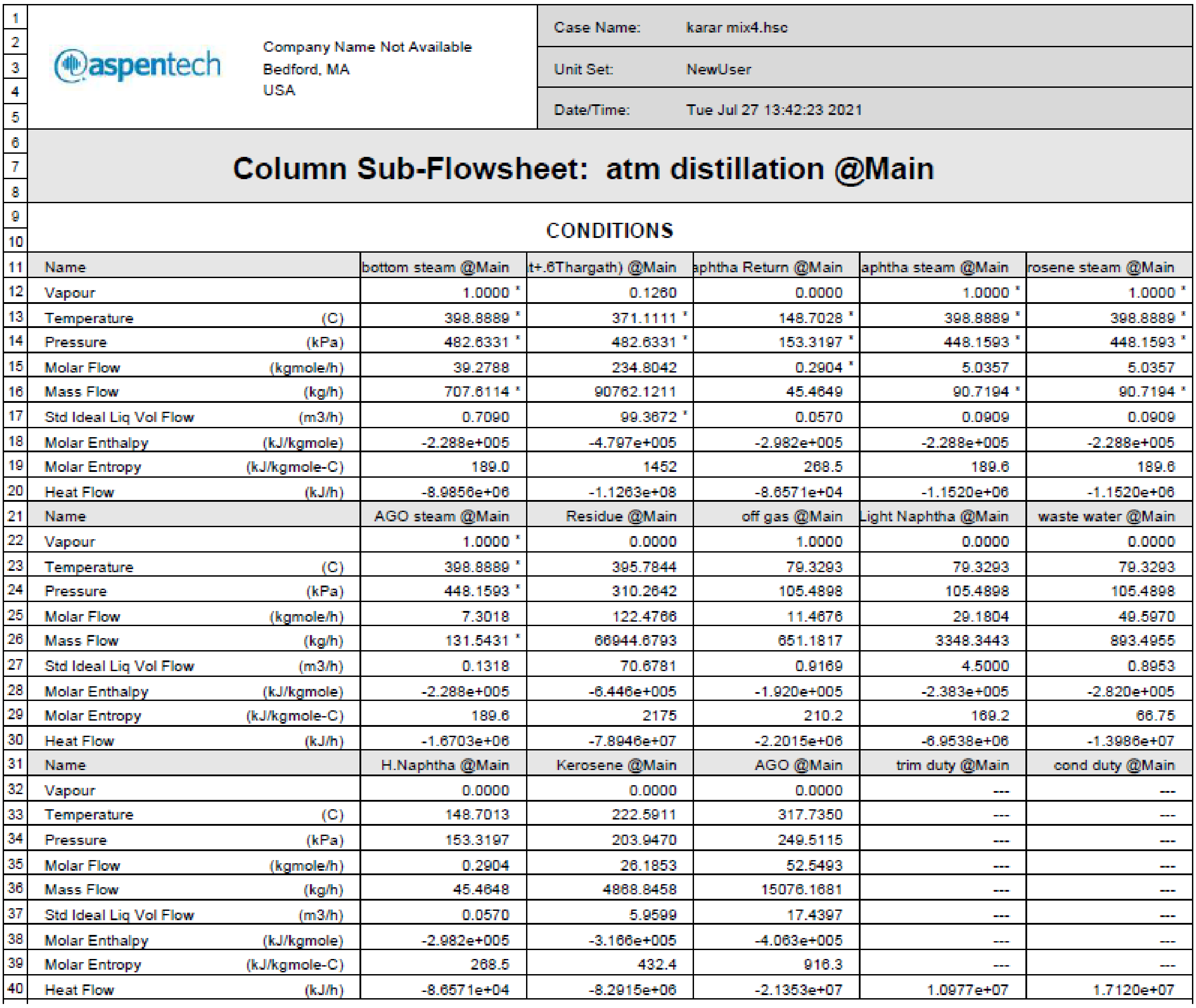
CONDITIONS

Name		bottom steam @Main	N+.133R+.2T) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2812	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	111.9740 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	260.4913	2.0298 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	89641.6517	36.5662	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.0366	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.223e+005	-2.794e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1288	73.75	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.1000e+08	-5.6712e+05	-1.1520e+06	-1.1520e+06

Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	373.0444	72.6111	72.6111	72.6111
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	122.9534	9.0461	46.4171	51.5040
Mass Flow	(kg/h)	131.5431 *	64478.1706	560.1415	4790.5855	927.8506
Std Ideal Liq Vol Flow	(m3/h)	0.1318	68.4013	0.8070	6.5860	0.9297
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.562e+005	-1.808e+005	-2.168e+005	-2.825e+005
Molar Entropy	(kJ/kgmole-C)	189.6	2023	203.6	140.1	65.24
Heat Flow	(kJ/h)	-1.6703e+06	-8.0686e+07	-1.6355e+06	-1.0061e+07	-1.4551e+07

Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	111.9721	199.9204	297.3911	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	2.0331	33.0795	54.1427	---	---
Mass Flow	(kg/h)	36.6261	5572.5722	14332.9150	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0367	6.9160	16.7498	---	---
Molar Enthalpy	(kJ/kgmole)	-2.794e+005	-2.977e+005	-3.915e+005	---	---
Molar Entropy	(kJ/kgmole-C)	73.75	357.4	804.8	---	---
Heat Flow	(kJ/h)	-5.6805e+05	-9.8464e+06	-2.1197e+07	3.5822e+06	1.8602e+07

Atmospheric Distillation: Mix 4 Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: karar mix4.hsc

Unit Set: NewUser

Date/Time: Tue Jul 27 13:42:23 2021

Column Sub-Flowsheet: atm distillation @Main

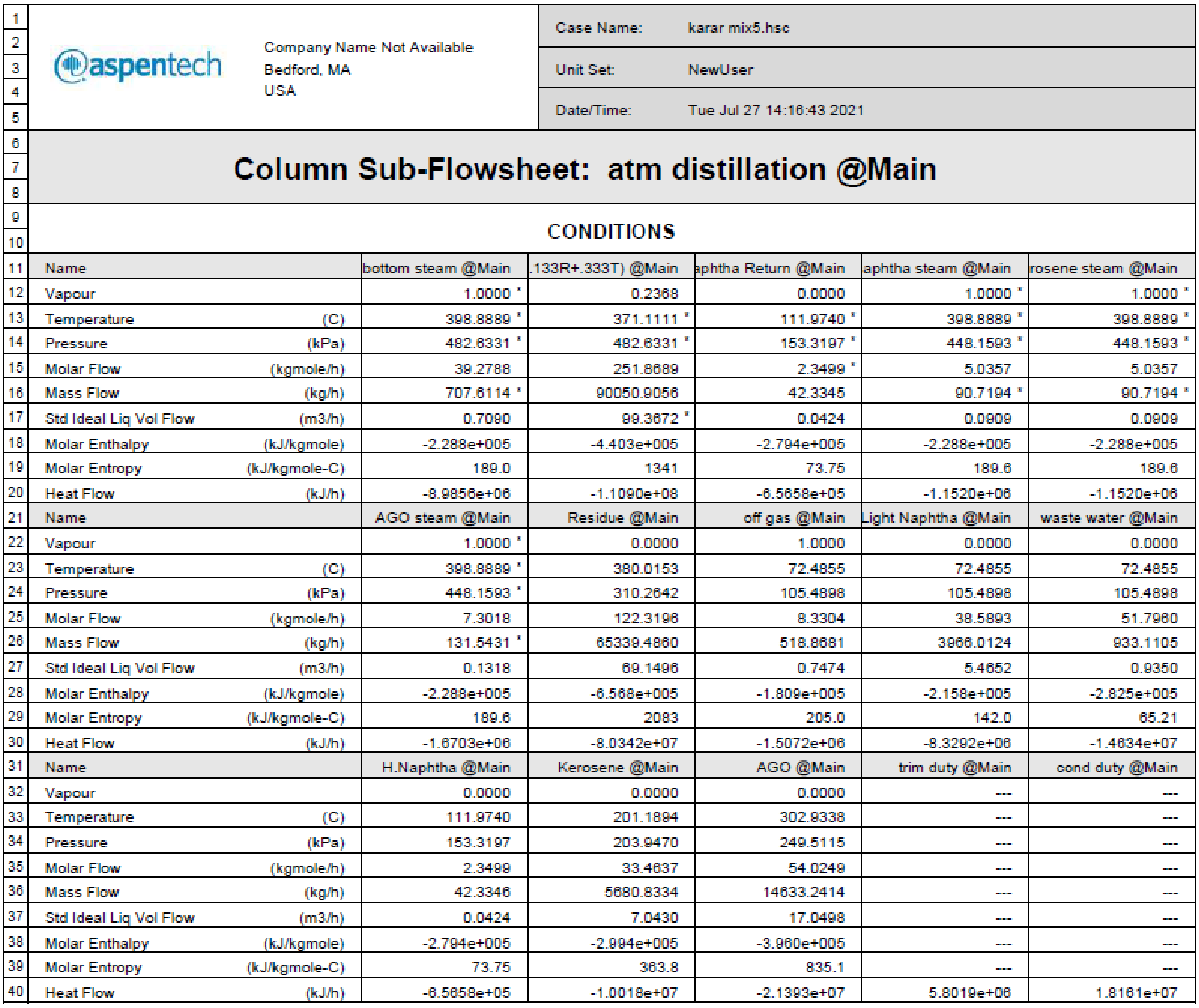
CONDITIONS

Name		bottom steam @Main	t+.6Thargath) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.1260	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	148.7028 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	234.8042	0.2904 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	90762.1211	45.4649	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.0570	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.797e+005	-2.982e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1452	268.5	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.1263e+08	-8.6571e+04	-1.1520e+06	-1.1520e+06

Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	395.7844	79.3293	79.3293	79.3293
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	122.4766	11.4676	29.1804	49.5970
Mass Flow	(kg/h)	131.5431 *	66944.6793	651.1817	3348.3443	893.4955
Std Ideal Liq Vol Flow	(m3/h)	0.1318	70.6781	0.9169	4.5000	0.8953
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.446e+005	-1.920e+005	-2.383e+005	-2.820e+005
Molar Entropy	(kJ/kgmole-C)	189.6	2175	210.2	169.2	66.75
Heat Flow	(kJ/h)	-1.6703e+06	-7.8946e+07	-2.2015e+06	-6.9538e+06	-1.3986e+07

Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	148.7013	222.5911	317.7350	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	0.2904	26.1853	52.5493	---	---
Mass Flow	(kg/h)	45.4648	4868.8458	15076.1681	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0570	5.9599	17.4397	---	---
Molar Enthalpy	(kJ/kgmole)	-2.982e+005	-3.166e+005	-4.063e+005	---	---
Molar Entropy	(kJ/kgmole-C)	268.5	432.4	916.3	---	---
Heat Flow	(kJ/h)	-8.6571e+04	-8.2915e+06	-2.1353e+07	1.0977e+07	1.7120e+07

Atmospheric Distillation: Mix 5 Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: karar mix5.hsc
Unit Set: NewUser
Date/Time: Tue Jul 27 14:16:43 2021

Column Sub-Flowsheet: atm distillation @Main

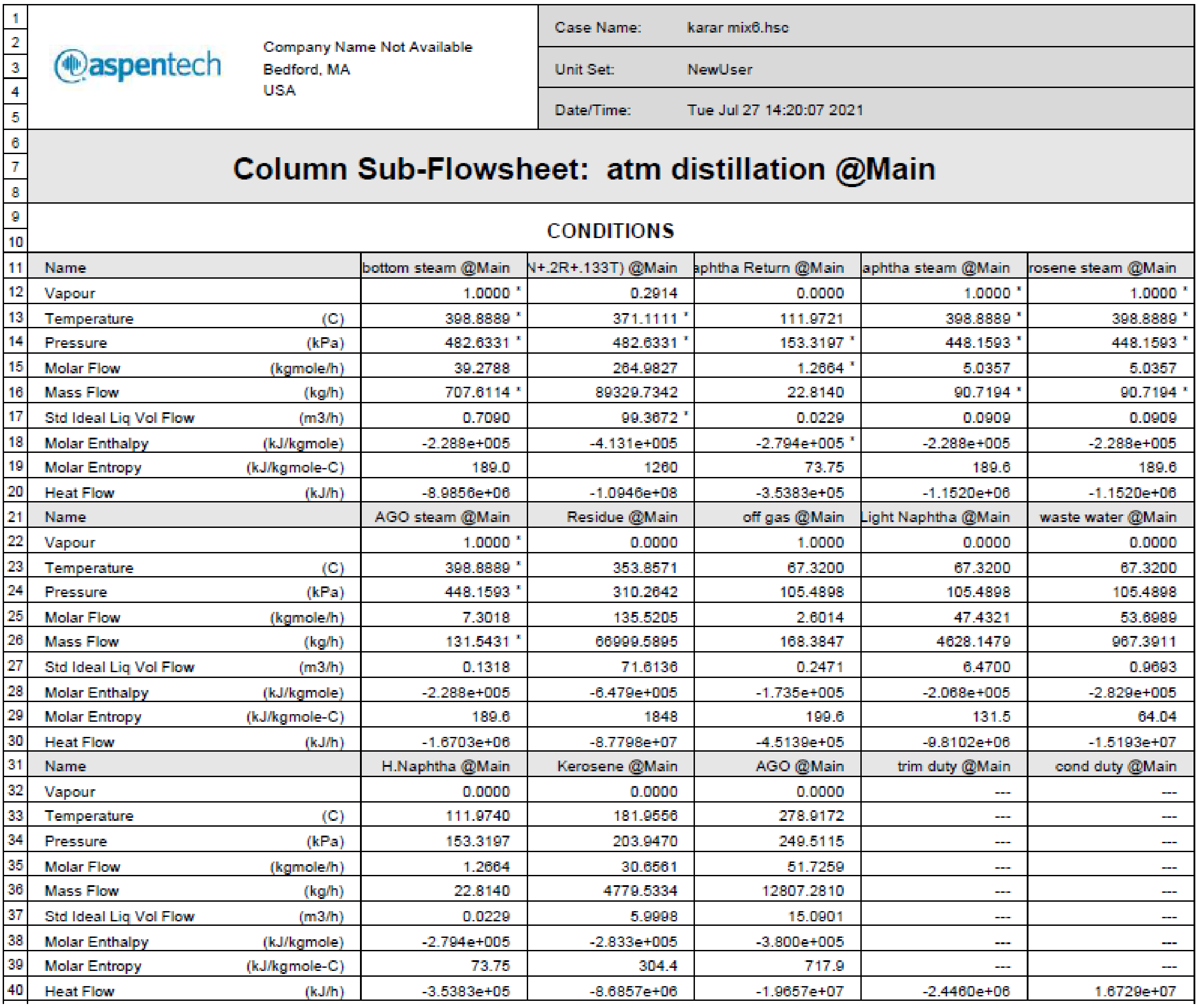
CONDITIONS

Name		bottom steam @Main	.133R+.333T) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.2368	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	111.9740 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	251.8689	2.3499 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	90050.9056	42.3345	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.0424	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.403e+005	-2.794e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1341	73.75	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.1090e+08	-6.5658e+05	-1.1520e+06	-1.1520e+06

Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	380.0153	72.4855	72.4855	72.4855
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	122.3196	8.3304	38.5893	51.7960
Mass Flow	(kg/h)	131.5431 *	65339.4860	518.8681	3966.0124	933.1105
Std Ideal Liq Vol Flow	(m3/h)	0.1318	69.1496	0.7474	5.4652	0.9350
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.568e+005	-1.809e+005	-2.158e+005	-2.825e+005
Molar Entropy	(kJ/kgmole-C)	189.6	2083	205.0	142.0	65.21
Heat Flow	(kJ/h)	-1.6703e+06	-8.0342e+07	-1.5072e+06	-8.3292e+06	-1.4634e+07

Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	111.9740	201.1894	302.9338	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	2.3499	33.4637	54.0249	---	---
Mass Flow	(kg/h)	42.3346	5680.8334	14633.2414	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.0424	7.0430	17.0498	---	---
Molar Enthalpy	(kJ/kgmole)	-2.794e+005	-2.994e+005	-3.960e+005	---	---
Molar Entropy	(kJ/kgmole-C)	73.75	363.8	835.1	---	---
Heat Flow	(kJ/h)	-6.5658e+05	-1.0018e+07	-2.1393e+07	5.8019e+06	1.8161e+07

Atmospheric Distillation: Mix 6 Conditions



Company Name Not Available
Bedford, MA
USA

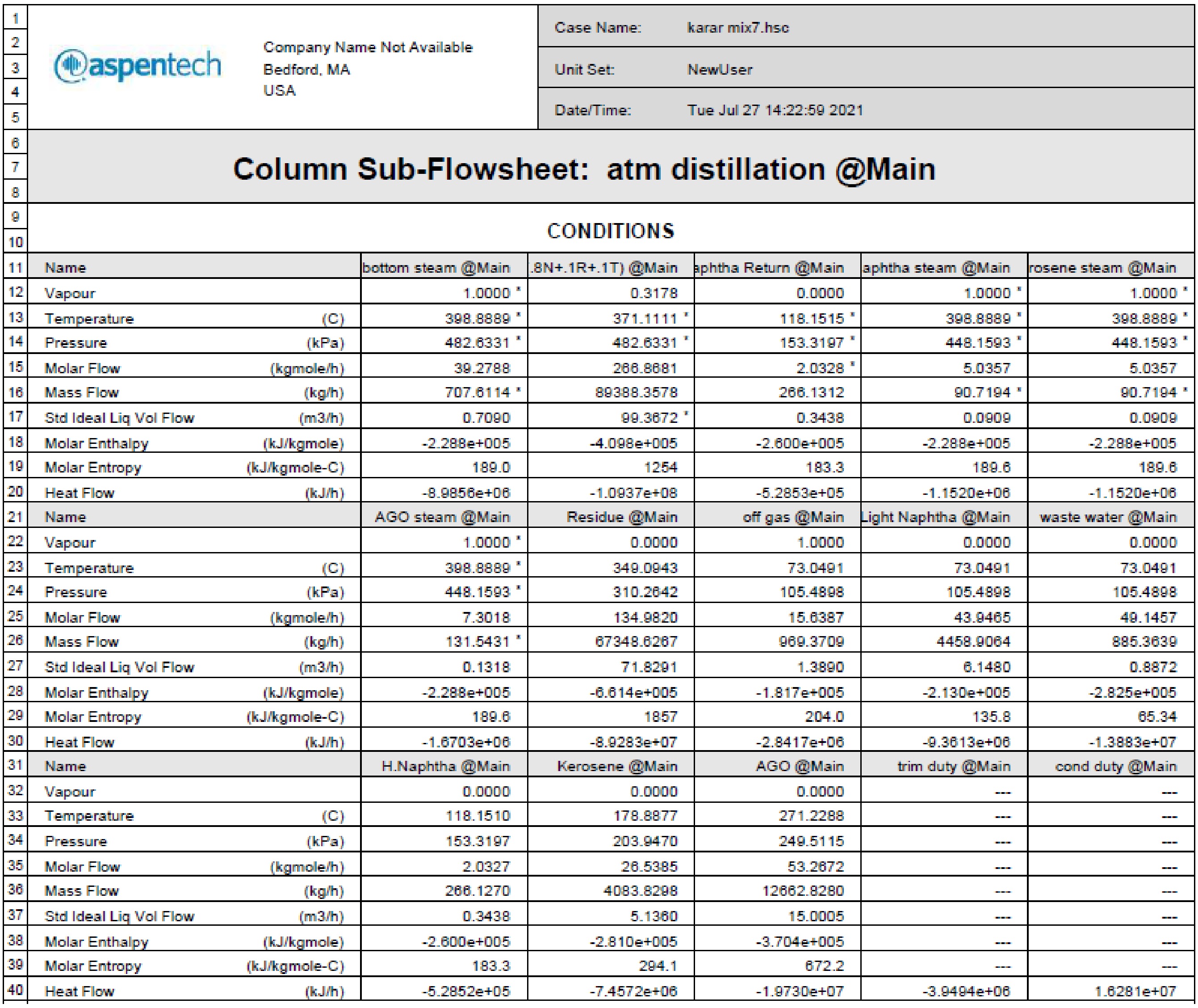
Case Name: karar mix6.hsc
Unit Set: NewUser
Date/Time: Tue Jul 27 14:20:07 2021

Column Sub-Flowsheet: atm distillation @Main

CONDITIONS

Name	bottom steam @Main	N+.2R+.133T) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour	1.0000 *	0.2914	0.0000	1.0000 *	1.0000 *
Temperature (C)	398.8889 *	371.1111 *	111.9721	398.8889 *	398.8889 *
Pressure (kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow (kgmole/h)	39.2788	264.9827	1.2664 *	5.0357	5.0357
Mass Flow (kg/h)	707.6114 *	89329.7342	22.8140	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow (m3/h)	0.7090	99.3672 *	0.0229	0.0909	0.0909
Molar Enthalpy (kJ/kgmole)	-2.288e+005	-4.131e+005	-2.794e+005 *	-2.288e+005	-2.288e+005
Molar Entropy (kJ/kgmole-C)	189.0	1260	73.75	189.6	189.6
Heat Flow (kJ/h)	-8.9856e+06	-1.0946e+08	-3.5383e+05	-1.1520e+06	-1.1520e+06
Name	AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour	1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature (C)	398.8889 *	353.8571	67.3200	67.3200	67.3200
Pressure (kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow (kgmole/h)	7.3018	135.5205	2.6014	47.4321	53.6989
Mass Flow (kg/h)	131.5431 *	66999.5895	168.3847	4628.1479	967.3911
Std Ideal Liq Vol Flow (m3/h)	0.1318	71.6136	0.2471	6.4700	0.9693
Molar Enthalpy (kJ/kgmole)	-2.288e+005	-6.479e+005	-1.735e+005	-2.068e+005	-2.829e+005
Molar Entropy (kJ/kgmole-C)	189.6	1848	199.6	131.5	64.04
Heat Flow (kJ/h)	-1.6703e+06	-8.7798e+07	-4.5139e+05	-9.8102e+06	-1.5193e+07
Name	H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour	0.0000	0.0000	0.0000	---	---
Temperature (C)	111.9740	181.9556	278.9172	---	---
Pressure (kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow (kgmole/h)	1.2664	30.6561	51.7259	---	---
Mass Flow (kg/h)	22.8140	4779.5334	12807.2810	---	---
Std Ideal Liq Vol Flow (m3/h)	0.0229	5.9998	15.0901	---	---
Molar Enthalpy (kJ/kgmole)	-2.794e+005	-2.833e+005	-3.800e+005	---	---
Molar Entropy (kJ/kgmole-C)	73.75	304.4	717.9	---	---
Heat Flow (kJ/h)	-3.5383e+05	-8.6857e+06	-1.9657e+07	-2.4460e+06	1.6729e+07

Atmospheric Distillation: Mix 7 Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: karar mix7.hsc

Unit Set: NewUser

Date/Time: Tue Jul 27 14:22:59 2021

Column Sub-Flowsheet: atm distillation @Main

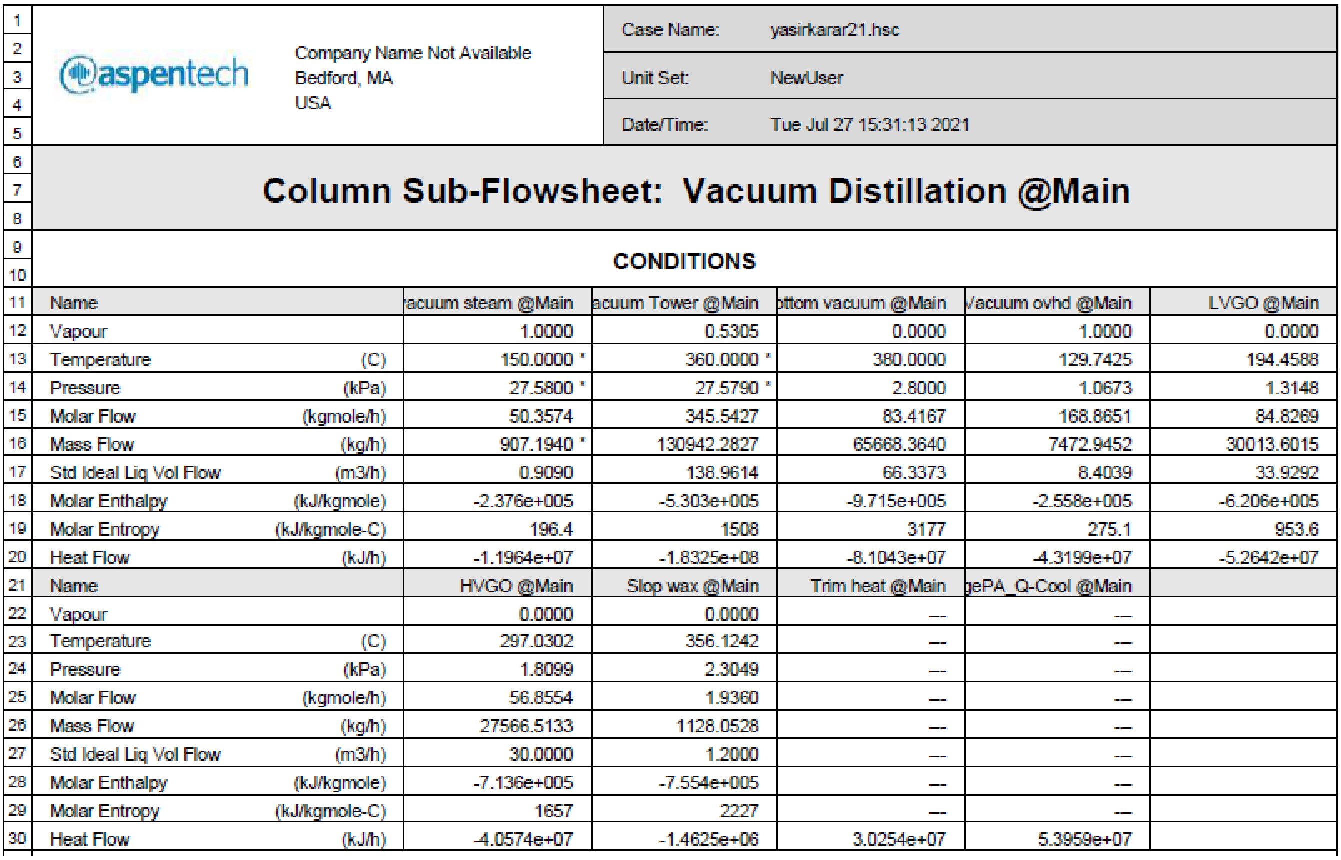
CONDITIONS

Name		bottom steam @Main	.8N+.1R+.1T) @Main	aphtha Return @Main	aphtha steam @Main	rosene steam @Main
Vapour		1.0000 *	0.3178	0.0000	1.0000 *	1.0000 *
Temperature	(C)	398.8889 *	371.1111 *	118.1515 *	398.8889 *	398.8889 *
Pressure	(kPa)	482.6331 *	482.6331 *	153.3197 *	448.1593 *	448.1593 *
Molar Flow	(kgmole/h)	39.2788	266.8681	2.0328 *	5.0357	5.0357
Mass Flow	(kg/h)	707.6114 *	89388.3578	266.1312	90.7194 *	90.7194 *
Std Ideal Liq Vol Flow	(m3/h)	0.7090	99.3672 *	0.3438	0.0909	0.0909
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-4.098e+005	-2.600e+005	-2.288e+005	-2.288e+005
Molar Entropy	(kJ/kgmole-C)	189.0	1254	183.3	189.6	189.6
Heat Flow	(kJ/h)	-8.9856e+06	-1.0937e+08	-5.2853e+05	-1.1520e+06	-1.1520e+06

Name		AGO steam @Main	Residue @Main	off gas @Main	Light Naphtha @Main	waste water @Main
Vapour		1.0000 *	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	398.8889 *	349.0943	73.0491	73.0491	73.0491
Pressure	(kPa)	448.1593 *	310.2642	105.4898	105.4898	105.4898
Molar Flow	(kgmole/h)	7.3018	134.9820	15.6387	43.9465	49.1457
Mass Flow	(kg/h)	131.5431 *	67348.6267	969.3709	4458.9064	885.3639
Std Ideal Liq Vol Flow	(m3/h)	0.1318	71.8291	1.3890	6.1480	0.8872
Molar Enthalpy	(kJ/kgmole)	-2.288e+005	-6.614e+005	-1.817e+005	-2.130e+005	-2.825e+005
Molar Entropy	(kJ/kgmole-C)	189.6	1857	204.0	135.8	65.34
Heat Flow	(kJ/h)	-1.6703e+06	-8.9283e+07	-2.8417e+06	-9.3613e+06	-1.3883e+07

Name		H.Naphtha @Main	Kerosene @Main	AGO @Main	trim duty @Main	cond duty @Main
Vapour		0.0000	0.0000	0.0000	---	---
Temperature	(C)	118.1510	178.8877	271.2288	---	---
Pressure	(kPa)	153.3197	203.9470	249.5115	---	---
Molar Flow	(kgmole/h)	2.0327	26.5385	53.2672	---	---
Mass Flow	(kg/h)	266.1270	4083.8298	12662.8280	---	---
Std Ideal Liq Vol Flow	(m3/h)	0.3438	5.1360	15.0005	---	---
Molar Enthalpy	(kJ/kgmole)	-2.600e+005	-2.810e+005	-3.704e+005	---	---
Molar Entropy	(kJ/kgmole-C)	183.3	294.1	672.2	---	---
Heat Flow	(kJ/h)	-5.2852e+05	-7.4572e+06	-1.9730e+07	-3.9494e+06	1.6281e+07

Vacuum Distillation: Nile Blend Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: yasirkarar21.hsc

Unit Set: NewUser

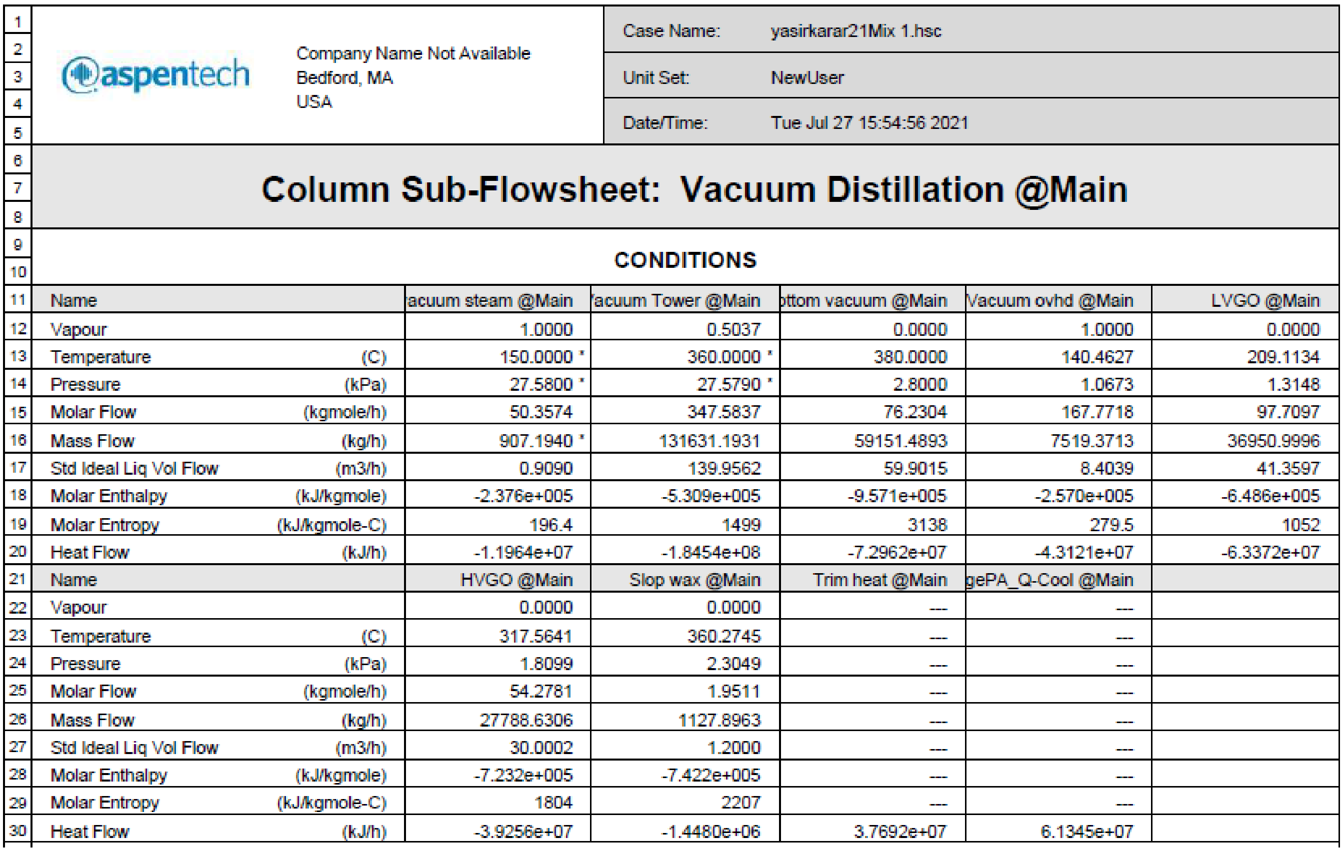
Date/Time: Tue Jul 27 15:31:13 2021

Column Sub-Flowsheet: Vacuum Distillation @Main

CONDITIONS

Name		acuum steam @Main	acuum Tower @Main	ottom vacuum @Main	Vacuum ovhd @Main	LVGO @Main
Vapour		1.0000	0.5305	0.0000	1.0000	0.0000
Temperature	(C)	150.0000 *	360.0000 *	380.0000	129.7425	194.4588
Pressure	(kPa)	27.5800 *	27.5790 *	2.8000	1.0673	1.3148
Molar Flow	(kgmole/h)	50.3574	345.5427	83.4167	168.8651	84.8269
Mass Flow	(kg/h)	907.1940 *	130942.2827	65668.3640	7472.9452	30013.6015
Std Ideal Liq Vol Flow	(m3/h)	0.9090	138.9614	66.3373	8.4039	33.9292
Molar Enthalpy	(kJ/kgmole)	-2.376e+005	-5.303e+005	-9.715e+005	-2.558e+005	-6.206e+005
Molar Entropy	(kJ/kgmole-C)	196.4	1508	3177	275.1	953.6
Heat Flow	(kJ/h)	-1.1964e+07	-1.8325e+08	-8.1043e+07	-4.3199e+07	-5.2642e+07
Name		**HVGO @Main**	**Slop wax @Main**	**Trim heat @Main**	**gePA_Q-Cool @Main**	
Vapour		0.0000	0.0000	---	---	
Temperature	(C)	297.0302	356.1242	---	---	
Pressure	(kPa)	1.8099	2.3049	---	---	
Molar Flow	(kgmole/h)	56.8554	1.9360	---	---	
Mass Flow	(kg/h)	27566.5133	1128.0528	---	---	
Std Ideal Liq Vol Flow	(m3/h)	30.0000	1.2000	---	---	
Molar Enthalpy	(kJ/kgmole)	-7.136e+005	-7.554e+005	---	---	
Molar Entropy	(kJ/kgmole-C)	1657	2227	---	---	
Heat Flow	(kJ/h)	-4.0574e+07	-1.4625e+06	3.0254e+07	5.3959e+07	

Vacuum Distillation: Mix 1 Conditions



Company Name Not Available
Bedford, MA
USA

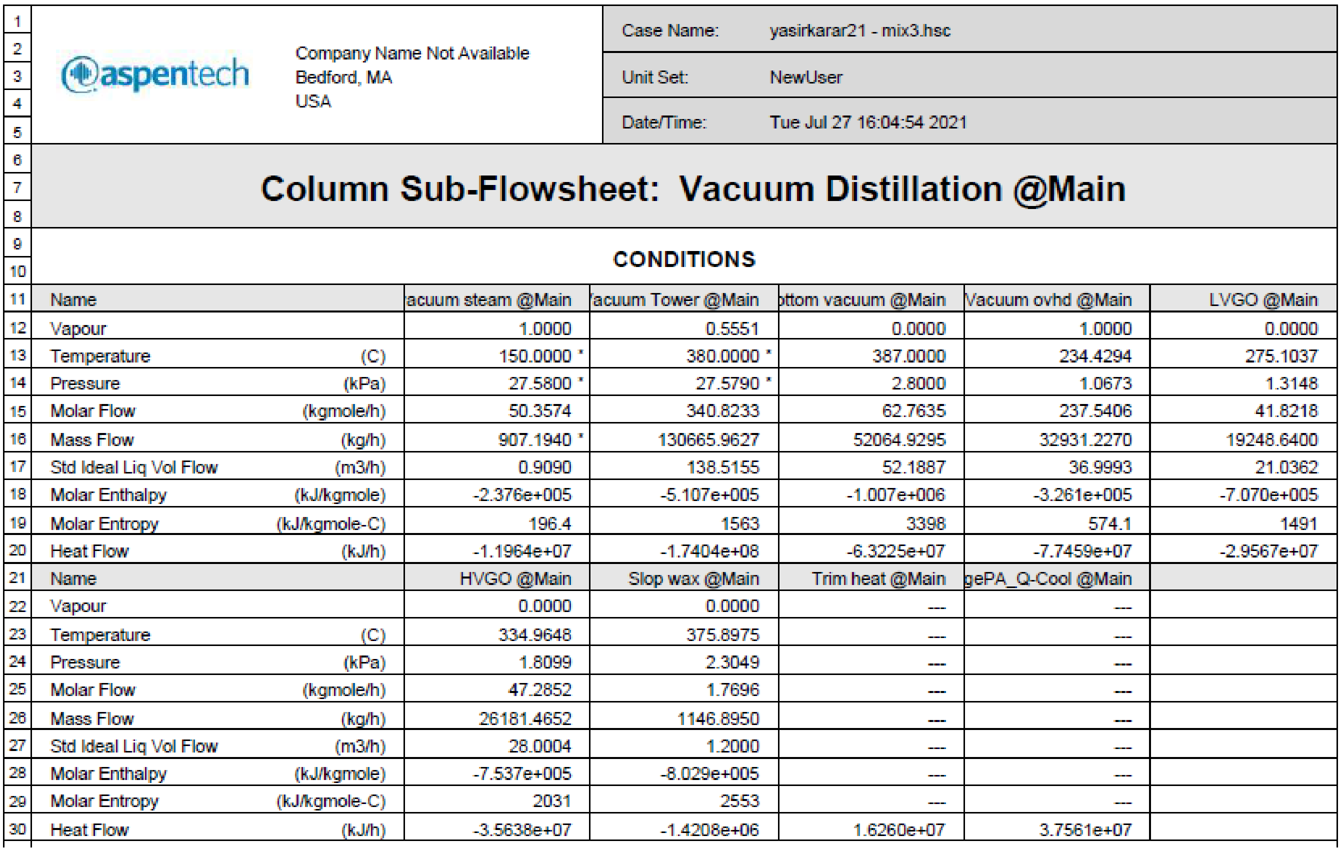
Case Name:	yasirkarar21Mix 1.hsc
Unit Set:	NewUser
Date/Time:	Tue Jul 27 15:54:56 2021

Column Sub-Flowsheet: Vacuum Distillation @Main

CONDITIONS

Name		acuum steam @Main	acuum Tower @Main	ottom vacuum @Main	Vacuum ovhd @Main	LVGO @Main
Vapour		1.0000	0.5037	0.0000	1.0000	0.0000
Temperature	(C)	150.0000 *	360.0000 *	380.0000	140.4627	209.1134
Pressure	(kPa)	27.5800 *	27.5790 *	2.8000	1.0673	1.3148
Molar Flow	(kgmole/h)	50.3574	347.5837	76.2304	167.7718	97.7097
Mass Flow	(kg/h)	907.1940 *	131631.1931	59151.4893	7519.3713	36950.9996
Std Ideal Liq Vol Flow	(m3/h)	0.9090	139.9562	59.9015	8.4039	41.3597
Molar Enthalpy	(kJ/kgmole)	-2.376e+005	-5.309e+005	-9.571e+005	-2.570e+005	-6.486e+005
Molar Entropy	(kJ/kgmole-C)	196.4	1499	3138	279.5	1052
Heat Flow	(kJ/h)	-1.1964e+07	-1.8454e+08	-7.2962e+07	-4.3121e+07	-6.3372e+07
Name		HVGO @Main	Slop wax @Main	Trim heat @Main	gePA_Q-Cool @Main	
Vapour		0.0000	0.0000	---	---	
Temperature	(C)	317.5641	360.2745	---	---	
Pressure	(kPa)	1.8099	2.3049	---	---	
Molar Flow	(kgmole/h)	54.2781	1.9511	---	---	
Mass Flow	(kg/h)	27788.6306	1127.8963	---	---	
Std Ideal Liq Vol Flow	(m3/h)	30.0002	1.2000	---	---	
Molar Enthalpy	(kJ/kgmole)	-7.232e+005	-7.422e+005	---	---	
Molar Entropy	(kJ/kgmole-C)	1804	2207	---	---	
Heat Flow	(kJ/h)	-3.9256e+07	-1.4480e+06	3.7692e+07	6.1345e+07	

Vacuum Distillation: Mix 3 Conditions



aspentech

Company Name Not Available
Bedford, MA
USA

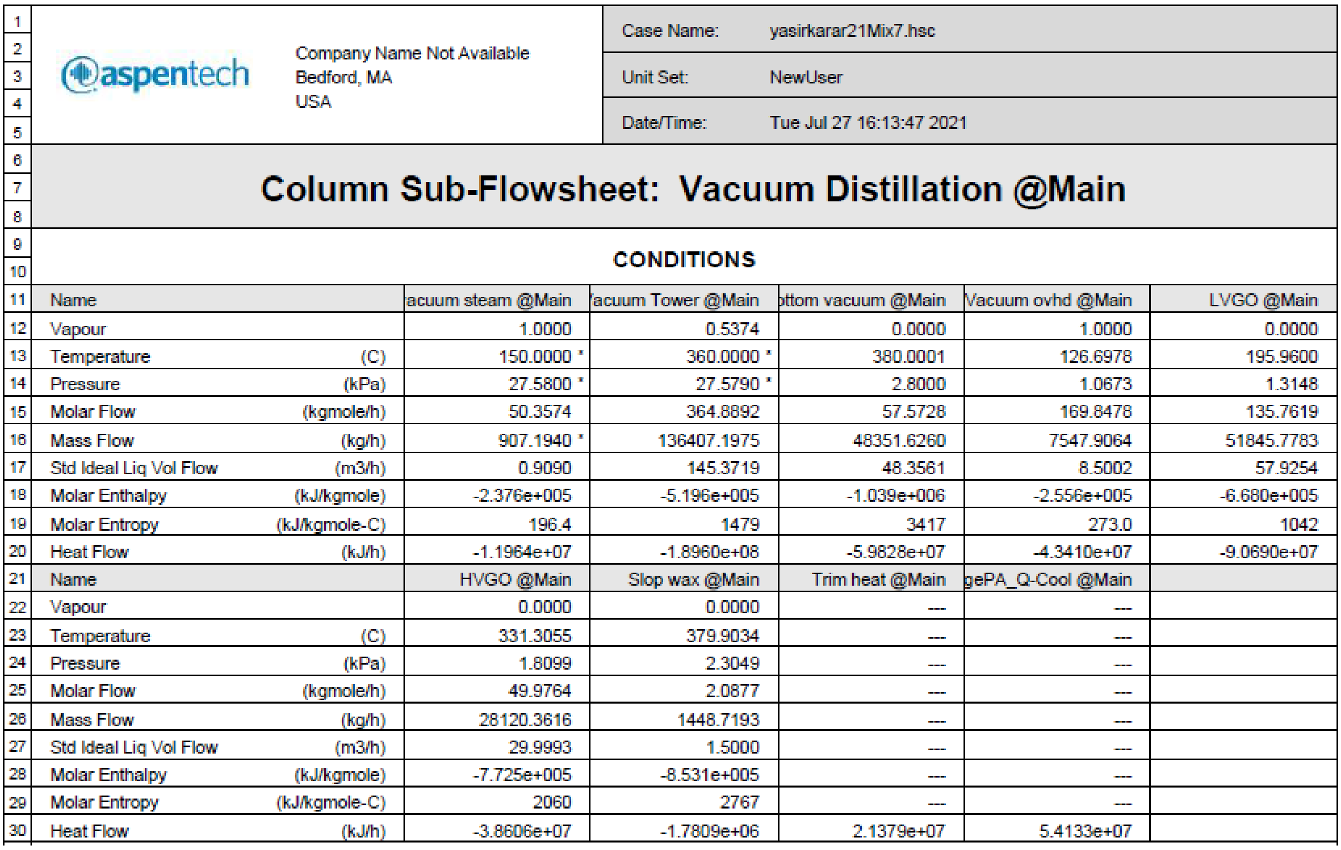
Case Name: yasirkarar21 - mix3.hsc
Unit Set: NewUser
Date/Time: Tue Jul 27 16:04:54 2021

Column Sub-Flowsheet: Vacuum Distillation @Main

CONDITIONS

Name		acuum steam @Main	acuum Tower @Main	ottom vacuum @Main	Vacuum ovhd @Main	LVGO @Main
Vapour		1.0000	0.5551	0.0000	1.0000	0.0000
Temperature	(C)	150.0000 *	380.0000 *	387.0000	234.4294	275.1037
Pressure	(kPa)	27.5800 *	27.5790 *	2.8000	1.0673	1.3148
Molar Flow	(kgmole/h)	50.3574	340.8233	62.7635	237.5406	41.8218
Mass Flow	(kg/h)	907.1940 *	130665.9627	52064.9295	32931.2270	19248.6400
Std Ideal Liq Vol Flow	(m3/h)	0.9090	138.5155	52.1887	36.9993	21.0362
Molar Enthalpy	(kJ/kgmole)	-2.376e+005	-5.107e+005	-1.007e+006	-3.261e+005	-7.070e+005
Molar Entropy	(kJ/kgmole-C)	196.4	1563	3398	574.1	1491
Heat Flow	(kJ/h)	-1.1964e+07	-1.7404e+08	-6.3225e+07	-7.7459e+07	-2.9567e+07
Name		HVGO @Main	Slop wax @Main	Trim heat @Main	gePA_Q-Cool @Main	
Vapour		0.0000	0.0000	---	---	
Temperature	(C)	334.9648	375.8975	---	---	
Pressure	(kPa)	1.8099	2.3049	---	---	
Molar Flow	(kgmole/h)	47.2852	1.7696	---	---	
Mass Flow	(kg/h)	26181.4652	1146.8950	---	---	
Std Ideal Liq Vol Flow	(m3/h)	28.0004	1.2000	---	---	
Molar Enthalpy	(kJ/kgmole)	-7.537e+005	-8.029e+005	---	---	
Molar Entropy	(kJ/kgmole-C)	2031	2553	---	---	
Heat Flow	(kJ/h)	-3.5638e+07	-1.4208e+06	1.6260e+07	3.7561e+07	

Vacuum Distillation: Mix 7 Conditions



Company Name Not Available
Bedford, MA
USA

Case Name: yasirkarar21Mix7.hsc

Unit Set: NewUser

Date/Time: Tue Jul 27 16:13:47 2021

Column Sub-Flowsheet: Vacuum Distillation @Main

CONDITIONS

Name		acuum steam @Main	acuum Tower @Main	ottom vacuum @Main	Vacuum ovhd @Main	LVGO @Main
Vapour		1.0000	0.5374	0.0000	1.0000	0.0000
Temperature	(C)	150.0000 *	360.0000 *	380.0001	126.6978	195.9600
Pressure	(kPa)	27.5800 *	27.5790 *	2.8000	1.0673	1.3148
Molar Flow	(kgmole/h)	50.3574	364.8892	57.5728	169.8478	135.7619
Mass Flow	(kg/h)	907.1940 *	136407.1975	48351.6260	7547.9064	51845.7783
Std Ideal Liq Vol Flow	(m3/h)	0.9090	145.3719	48.3561	8.5002	57.9254
Molar Enthalpy	(kJ/kgmole)	-2.376e+005	-5.196e+005	-1.039e+006	-2.556e+005	-6.680e+005
Molar Entropy	(kJ/kgmole-C)	196.4	1479	3417	273.0	1042
Heat Flow	(kJ/h)	-1.1964e+07	-1.8960e+08	-5.9828e+07	-4.3410e+07	-9.0690e+07
Name		HVGO @Main	Slop wax @Main	Trim heat @Main	gePA_Q-Cool @Main	
Vapour		0.0000	0.0000	---	---	
Temperature	(C)	331.3055	379.9034	---	---	
Pressure	(kPa)	1.8099	2.3049	---	---	
Molar Flow	(kgmole/h)	49.9764	2.0877	---	---	
Mass Flow	(kg/h)	28120.3616	1448.7193	---	---	
Std Ideal Liq Vol Flow	(m3/h)	29.9993	1.5000	---	---	
Molar Enthalpy	(kJ/kgmole)	-7.725e+005	-8.531e+005	---	---	
Molar Entropy	(kJ/kgmole-C)	2060	2767	---	---	
Heat Flow	(kJ/h)	-3.8606e+07	-1.7809e+06	2.1379e+07	5.4133e+07	