PROCESS SIMULATION AND OPTIMIZATION OF A CRUDE OIL DISTILLATION UNIT

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MSc THESIS

Chemical Engineering Department

Supervisor: Prof. Dr. Fethi KAMISLI

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Ari Abdulqader Ahmed

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ABSTRACT

In the present study simulation and optimization of crude distillation Unit of Erbil Refinery by HYSYS simulation program has been carried out with the aim of increasing the efficiency of this plant.

Different alternatives were evaluated with using the different operational variables, such as the temperature of the kerosene removal plate, the top temperature of the distilling tower, the stripping steam flow, the transfer temperature and the top pressure of the tower. In order to determine the influence each variable on the performance of the unit, one of the variables was changed while keeping the rest of variables to be constant. It was determined that the yield of kerosene increased to be 5.98 % by varying the temperature of the kerosene removal plate from 202 ˚C to 206 ˚C. This variation causes an increase in sulfur content and a decrease in diesel cutting since a part of diesel converts into kerosene. It was observed that reducing the cap temperature from 153 ˚C to 150 ˚C helped to increase low rate of kerosene from 3.2 m^3/h to 4.7 m^3/h while flow rate of naphtha decreased from 32 m^3/h to 30 m^3/h since some of naphtha became a part of kerosene that decreased sulfur content in kerosene. It was demonstrated that the increases of both the stripping steam flow in the bottom of the distilling tower and the tower feeding temperature increase flow rates of both the kerosene and the naphtha cuts from 3.2 m^3/h to 7 m^3/h and from 32 m^3/h to 35 m^3/h, respectively. It was noted that the adjustment generates the greatest impact on the yields of the products with the highest commercial value without generating additional costs. In other words, the reduction of the pressure from 1.7 kg/cm^2 to 1.5 kg/cm^2 is enough to accomplish this. It was observed that the optimum operation conditions determined by the simulation increased the efficiency of the plant since the yields of both kerosene and naphtha increased to be 5.98 % with the specified quality and 29.91% with the specified quality, respectively.

As a result, there are increases in the recoveries of kerosene, naphtha and diesel, which are highly demanded since they are greater commercially-valuable products than the residue. An increase in commercially-valuable products (lighter cuts) than the residue is of great importance for this study.

Keywords: HYSYS, Simulation, Optimization, Crude distillation unit.

ÖZET

Ham Petrol Distilasyon Ünitesinin Süreç Simülasyonu ve Optimizasyonu

Bu çalışmada, işletmenin verimini arttırmak amacıyla Erbil rafinerisinin ham petrol distilasyon ünitesinin simülasyonu ve optimizasyonu HYSYS simülasyon programı ile yapıldı.

Farklı alternatifler; gazyağı uzaklaştırma plaka sıcaklığı, sıyırıcı buhar debisi, transfer sıcaklığı ve kule tepe basıncı gibi farklı işletme değişkenler kullanılarak denendi. Birimin performansı üzerinde her bir değişkenin etkisini saptamak için diğer değişkenler sabit tutularak değişkenlerden biri değiştirildi. Gazyağı uzaklaştırma plakasının sıcaklığı 202 ˚C ’den 206 ˚C ’ye değişmesiyle gazyağı verimi % 5.98 arttığı belirlendi. Bu değişim sülfür içeriğinde artmaya ve dizel payında azalmaya neden olmaktadır. Çünkü dizelin bir kısmı gazyağına dönüşmektedir. Tepe sıcaklığının 153 ˚C ’den 150 ˚C ’ye düşürülmesi; nafta debisini 32 m^3/sa’den 30 m^3/sa’te düşürürken, gazyağı debisini 3.2 m^3/sa’den 4.7 m^3/sa’te çıkardığı gözlendi. Çünkü naftanın bir kısmı gazyağına dönüşmüş ve bu, gazyağı içerinde sülfür içeriğini düşürmüştür. Hem distilasyon kulesinin tabanında sıyırma buharının debisinin hem de kule besleme sıcaklığının artırılmaları; hem gazyağı hem de nafta debilerini sırasıyla 3.2 m^3/sa’den 7.0 m^3/sa’te ve 32 m^3/sa’den 35 m^3/sa’te çıkardığı gözlendi. Ayarlamaların ilave masraf çıkarmadan en yüksek ticari değere sahip olan ürünlerin verimleri üzerinde en büyük etkiyi oluşturdukları not edildi. Diğer bir ifadeyle, bu başarmak için basıncın 1.7 kg/cm^2’den 1.5 kg/cm^2’ye düşürülmesi yeterlidir. Simülasyonla belirlenen optimum işletme şartları işletmenin verimini arttırdığı gözlendi. Çünkü belirlenmiş kalitede (kalite düşürülmeden) hem gazyağı hem de naftanın verimleri sırasıyla % 5.98 ve % 29.91 artmıştır.

Sonuç olarak, gazyağı, nafta ve dizel’ in geri kazanımları artmaktadır. Bu, oldukça talep edilen bir durumdur. Çünkü bu ürünler ticari olarak yağyakıt/asfalttan (residue) daha değerlidirler. Yağyakıt/asfalttansa ticari olarak daha değerli olan ürünlerdeki (daha hafif ürünler) artım, bu çalışma için büyük öneme sahiptir.

Anahtar kelimeler: HYSYS, Simülasyon, Optimizasyon, Distilasyon ünitesi.

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SOME SYMBOLS AND ABBREVIATIONS

Some symbols and abbreviations used in this study are presented below, along with descriptions.

Symbols	**Description**
Conc.	Concentration
g	Gram
kg	Kilogram
m	Meter
cm	Centimeter
mm	Millimeter
l	Liter
ml	Milliliter
˚C	Celsius
%	Percentage
ppm	Part per million
PID	Piping and Instrumentation Diagram
PFD	Process Flow Diagram
ppt	part per thousand
PTB	part per thousand barrel
“	symbol for second in time
‘	symbol for minute in time
Vol	volume
W/V	water per volume
V/V	volume per volume
ASTM	American Society for Testing and Materials
APC	Advanced Process Control
API	American Petroleum Institute
SPS	Software Package Process Simulation

1. INTRODUCTION

The history of process simulation strongly has to do with the occurrence of computer science, hardware and programming languages. The early performance that easy implementations of partial aspects of chemical processes operations were introduced in the 1970s once the appropriate hardware and software package (being the main FORTRAN programming language) have been made available. The modeling of the chemical properties began to abound especially when the cube equation of the States and Antoine´s equation were developed in the nineteenth century (Chryssolouris, 2005).

1.1 Process Simulation Mode

There are two types of forms:

(a) Stable State or Stable conditions
(b) Simulation of Dynamic Processes

Initially, the simulation of processes used to simulate steady state processes. Steady-State Models perform a mass and energy balances of stationery process (a process in a State of Equilibrium) and do not depend on time.

Dynamic Simulation is an extension of the simulation of processes in stable condition in which by-products such as mass and energy can be accumulated. The emergence of dynamic simulation implies that actual real-time processes time-dependent description, prediction and management have become attainable. This includes schema of startup and shutdown of the plant, changing conditions along a reaction, bottlenecks, thermal changes, etc...

The dynamic simulations require time calculation high and are mathematically much more advanced than a simulation of stable state. This is a simulation of steady state continuous multiple (based on a passage of time set) with parameters that are constantly changing.

Dynamic simulation can be used in each of the Web and offline forms. The Web case being the Prophetic management model, where the results of the period simulation are accustomed, predicts the changes that can occur for a change of input effect and, therefore, the management parameters are optimized for the results. Offline process simulation can be used in planning, solution of problems and improving the process plant because the physical phenomenon of the case studies can be used to evaluate the impacts of process modifications. Dynamic simulation is used additionally for operator training.

Modeling of simulation of software processes: As in all simulations, the simulation of processes of software packages (SPS) is the numerical analysis of a mathematical model that copies the behavior of the process development. SPS will model the dynamic nature of the process development and will handle the uncertainty and the randomness inherent in it (Robertson and Palazoglu, 2011).

1.2 Software Package Process Simulation

The simulation of the process with software begins with the distinction of a question that everyone wants to answer. The question may be, for example, associated with the evaluation of other, incorporating a replacement note within the process development with the software package. Such changes within the current development process are expensive and if the results of the change do not seem to be positive, the implications can be disastrous for the organization. Therefore, simulation of a process that has a tendency to obtain an initial assessment of these changes in the model instead of a vigorous development of project. One can choose an acceptable scope of the process. A simulation approach can be chosen to model the process of the event. Such a model is then a brand of exploitation of empirical information, so regular behavior simulation is based mainly on research. One can find a detailed description of each general briefs on Balci's work (2012) and for the simulation of the process of the software package will be a comprehensive summary in the book of (Roses, 2012).

Chemical engineering can be defined from many different aspects. Nevertheless, all the scientists and professionals agree in that the process is the center of itself. To do a distinction of any another discipline, the role of the chemical engineering might be defined by the intention of developing, of designing, of constructing, of controlling, of optimizing and of managing any process that implies physical changes and / or chemical and of doing this profitable process without violating the environmental balance. The simulation of processes as a discipline uses mathematical models as a basis for analysis, prediction, testing, and detection of a behavior of the process.

Process simulation is there to increase the level of knowledge for a process and chemical engineering in general. Then, when these two concepts are combined, one can consider the chemical engineering as a discipline that defines how to develop the process and the simulation as the tool that helps us to explore the options. The chemical engineering needs to know how to design the process, while the chemical engineers use the simulation to explore all the options of process design and define the ideal one.

The process simulation is applied at present in almost all the disciplines of chemical engineering and engineering in general. It is the inevitable part of the different disciplines such as the design of the process, the investigation and the development, the planning of the production, the optimization, the training and the education and the decision-making for a process. It can be said that the process simulation is one of the most important parts in the engineering. A wide palette of solutions of simulation is given in Figure 1.1.

Figure 1.1 Pallet of different area of simulation

1.3 Stages for Develop a Process Simulation using Specialized Software

Develop a chemical process simulation needs to apply the following procedure

a) Process design.
b) Research and development of process.
c) Production and planning.
d) Running simulation and corrections.
e) Optimization.
f) Taking decisions.

Process Design

Process design represents one of the traditional simulation and mathematical modeling applications. Process synthesis and process design use stationary state models to define the process flow diagram accompanied by heat balance and material. The objectives of the process design are to find the best process diagram and the best conditions of design. This can be a complex task that needs to explore large number of options and is not possible without the use of mathematical modeling and simulation of processes.

Research and Development of Process

The chemical engineering is as a source of challenges that produces an ongoing inspiration for investigators in their projects. No research project is possible without a certain quantity of mathematical models and simulation of processes involved. Therefore, it can minimize the quantity of experimental investigation. There are certain parts of the process that continuously need assessment and improvement. Sections of reactors are very often particular part of the process, especially if the catalytic converter is involved in the reactions. With this end, a continuous supervision of the yield is realized to change the conditions of the reactor at the appropriate time. Engineers who work with research and development are involved in detailed mathematical models that include a large number of physical properties and thermodynamic to help them assess the current conditions or improved process.

Production and Planning

The planning and programming of production accompanied by economic calculations represent an important discipline that is placing a chemical process or manages on the market. Ones the process is executed; its profitability turns into one of the most important tasks for a chemical engineer. The profitability of the process is being scanned and is defined through the planning of production and the models that are used to provide the answers to questions about how to define the production and the ideal functioning.

Market change, change in feeds and products need constant assessment to ensure profitability. Mathematical models are used to simulate all possibilities as a guide on the way to the optimal solution. This is done to help management to take the correct decisions.

Running Simulation and Correction

The dynamic simulation analyzes the ideal functioning of the process, safety, the environmental limitations and the controllability to help define strategies of control, objectives and control parameters.

The dynamic simulation is used for the first time during the design phase of the process to help define control strategies. When the process is in functioning, it is used to analyze, to test and to optimize the operation conditions. This type of analysis can give answers about the bottlenecks of the process and how to solve them. As "Time" is included as a variable, dynamic simulation centers on control problems in which steady state simulation cannot meet requirements.

Optimization

The dynamic models allow the chemical engineers to execute continuously the unit with a strategy of definite optimization, transforming the knowledge of the process into the form of the mathematical model hidden within the control algorithm, called Advanced Process Control (APC).

This approach provides to the engineers and operators the capacity and the operators of almost run the unit like operating the aircraft on autopilot, caring constantly of the economic benefits (Zhang, 2000).

Taking Decisions

The process of decision-making endorsed by different types of calculations, models and simulations is much more efficient than one based in assumptions. A complete formulation exists of how different models can support the decision making process so that it is less infuriating and difficult.

1.4 Study Area

This study was developed in Erbil Refinery, motivated in trying to support solving problems about crude oil yield to determine operational parameters to improve to maximize production (volume produced of) some distilled fraction like gasoline and kerosene.

Considering every action being taken it is very important to highlight that, in the Khormala region, a project of great magnitude associated with oil and gas is being carried out, which is producing around three hundred thousand barrels of crude oil per day. However, there is one project that is currently being carried out, which stands out for being crude oil extraction and its purification process to export to Erbil KAR known as the Central Processing Station Central Process Station receives crude oil and gas from every Oil Well. This complex is responsible for the separation of impurities that are entrained in the crude and gas (mud and undesirable condensates) with a production of heavy acid gases (high content of hydrogen sulfide-H_2S and carbon dioxide CO_2) and desalinated crude ready to be refined as the block diagram shown in Figure 1.2.

In the extraction process it is common receiving water assisted oil and salts from respective location, therefore, CPS considers three (03) stations: south, north, and middle stations (reservoirs efficiency). During the operation process, it has been demonstrated existence of amount of salts within distillation of fossil fuels. In order to reduce this amount of salt in the oil combined with water, several desalination plants were installed in the process to halt such emulsion and tip out salts specifically, Khormala Oil and Gas Project (KOG) is located in the southwest of Erbil-Kurdistan, 65 kilometers from the city. KAR Group, a Kurdistan based Oil Service Company, runs operations in the Oil Field known as Khormala which has 60 oil wells allocated around the wells area identified as north, middle and south stations.

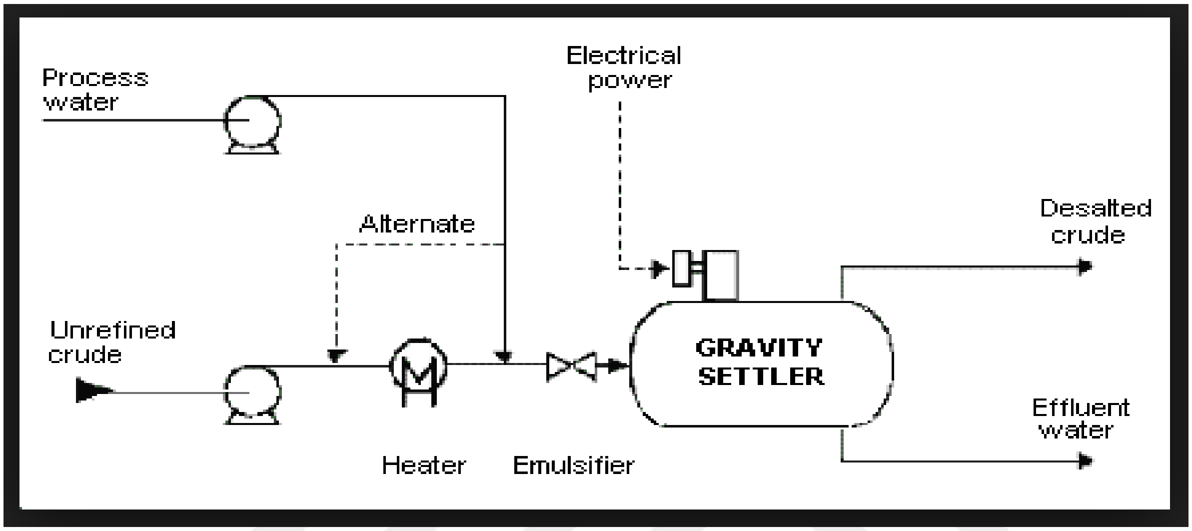


Figure 1.2 Desalter scheme (own, 2017)

The overall scheme of extraction of crude oil and received by Erbil Refinery is shown in Figure1.3.

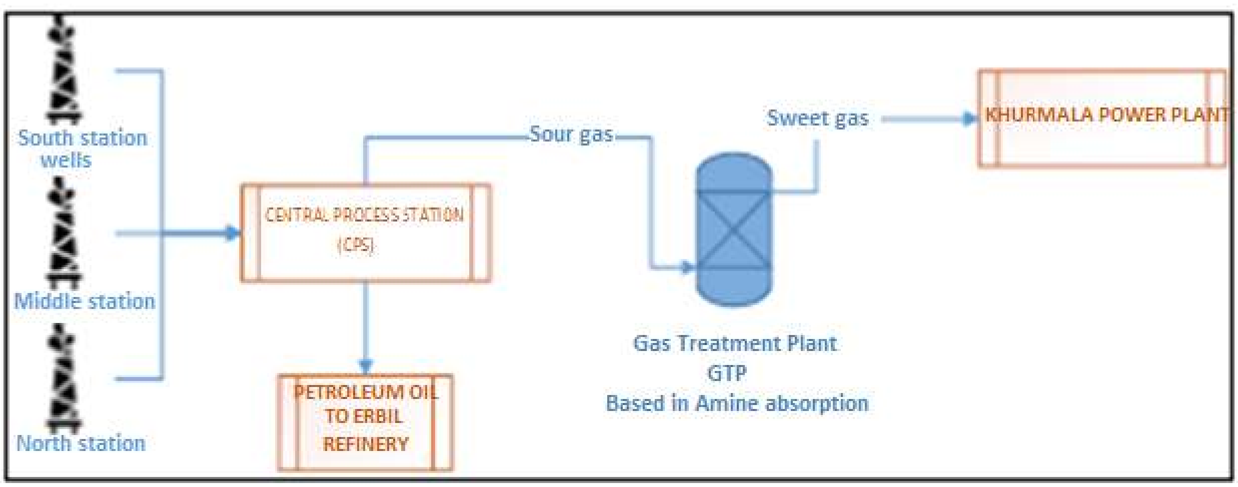


Figure 1.3 Block diagram configuration of Khormala oil and gas project (own, 2017)

The atmospheric distillation plant has the conformation specified in appendix A "Process and Flow Diagrams" specifies 2 Process Flow Diagrams (PF-102 and PF-101). Distillation unit has a design capacity of 20,000 BSPD (or Barrel Standard Per Day–BPD). However, nowadays at normal conditions, Atmospheric Distillation Unit (ADU) is processing around 117 m^3/hr, that means it is operated at 87.5 % of design capacity. The yields based on distilled fractions are calculated based on ASTM D86 “Unit - Operational Report” are presented in Table 1.1.

Table 1.1 Normal yields of ADU (Own, 2017)

Fraction distilled	Flow (m^3/hr)	Volumetric yield (%)
Naphtha	32	27.36
Kerosene	3.2	2.74
Diesel	16.4	14.02
Atmospheric Gas Oil (AGO)	1.5	1.28
Atmospheric Residue (ATB)	63.87	54.6
Total	116.97	100.0

In another order of ideas, the report related to Khormala Crude Oil Evaluation (ASTM D 2892) shows the performance of crude oil as reference at distillation process. It can be seen that in this Crude Oil Assay Report, the kerosene yield obtained by standard Assay ASTM D 5236 and D 2892 using API algorithm is 8.5 % vol. This value was obtained using Maxwell and Bonnel interpolating kerosene gross yield between fractions in (160 – 280) ˚C and then, making a rigorous tuning calculation for theoretical yield for actual interval (151 – 202) ˚C is 6.2 % vol.

This reference value gives the opportunity for improving ADU in the way to obtain more volume of kerosene stream at actual conditions. Therefore, the opportunity to improve the performance in the cutoff of kerosene in the actuals processing conditions will be near 3 % vol that is almost double of actual yield.

The comparison of actual and theoretical yield of kerosene is given in Table 1.2 in which ADU is seen to be improved at kerosene fraction at maximum of 47 % from actual yield. It can be increased total volume flow rate of production to be increased from range of (2.7 – 3.2) m^3/h up to (3.97 – 4.7) m^3/h.

According to information presented in Table 1.2, ADU in Erbil Refinery must be optimized by applying a chemical process simulation to maximize the production of kerosene fraction by increasing the produced flow rate.

Table 1.2 Comparison of actual and theoretical yield of kerosene production at ADU (Own, 2017)

Kerosene cutoff stream (151 – 202) ˚C		
Yield from Diesel Side	8% vol	Average: 8.5%
Yield from Naphtha Side	9% vol	
Maxwell Correction Factor	0,8	
Edminster Correction (50%)	0.76	
Reference Kerosene Yield	5,168%	
Difference on kerosene yields (Reference – Actual)	2.428%	
% of Difference	47%	

This could be made if crude oil is submitted in some way to adjust operational parameters by sensitivity study of pressures, flows and temperatures at T-101 and Kerosene Splitter T-104 with related heat exchangers. In this study, the interest is to determine the suitable conditions to achieve a maximum volume flow rate of outlet kerosene stream with good quality from ADU.

It is important to mention that trying to achieve this accomplishment directly without any previous study, the undesirable results and unsafe condition and environmental risks can be taken place in the operational complex. Changing real conditions have limitations like:

a) Distillation Unit cannot be submitted to a test run for researching or making tests to achieve those conditions because is an essential unit that shall be working continuously. Using this unit in untested operational new conditions, will distortion downstream processes and produce all products out of quality specifications.
b) Technology and operational limitations come from different equipments.

With this in mind, this research makes a proposal to determine operational conditions, efficiency, performance and balances using specialized computational tools namely HYSYS ASPEN simulation software. The fundamental reason why HYSYS ASPEN V8.8 was selected is of reliability about algorithm of calculations, conversions of results to real process. and most important, the design and validation of specification over crude oil unit (ADU) plant design and construction were done by original manufacturer (VENTEC ENGINEERS INC) over same HYSYS software (See Appendix A – Process Flow Diagram and Material Balance Report for

Crude Oil System). That is why to use same baseline calculations protocols HYSYS ASPEN was selected and used.

The layout of complex is seen in Figure 1.4 and it will be explained in detail in next section why computational tools must be used instead of “on run study conditions” cannot be done in present study.

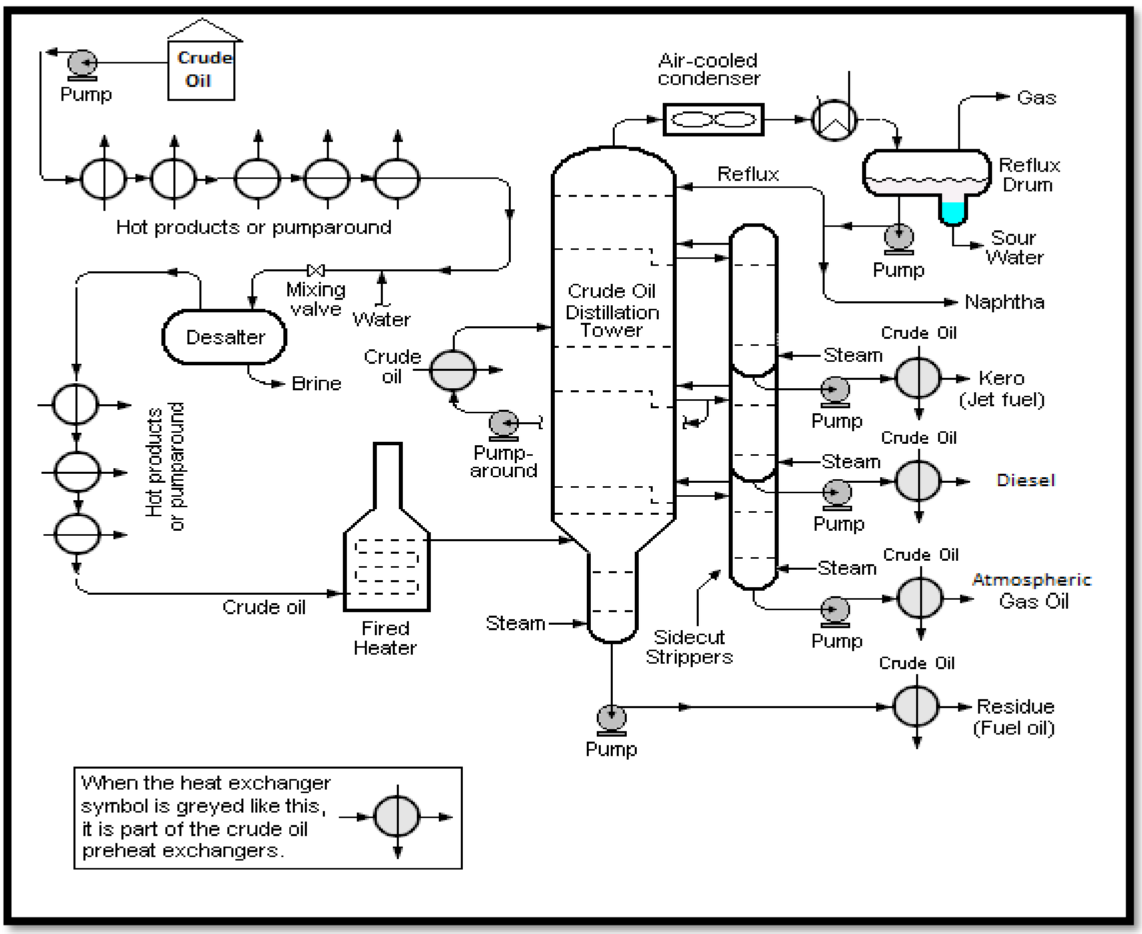


Figure 1.4 Block diagram of Erbil refinery crude distillation unit

Based on the explained facts in previous paragraphs, ADU needs to be improved. In order to do this, might be, the most suitable way is to apply computational tools for determining its performance, which motivates this study to achieve above objectives. The main aim of this project is to simulate a Crude Distillation Unit with an inlet crude composition of crude oil terminal act as the receiving facilities to obtain stabilized crude oil distillation unit with maximum approachable conditions for refining.

1.5 Goal and Objectives of Research

1.5.1 Main Goal:

Simulation and Optimizing of Distillation Process in Erbil Refinery

1.5.2 Specific Objectives:

- Identify and determine main equipments and parameters getting the field ranges of working pressures, temperatures, compositions and flows in normal operations.
- Simulate distillation process in a rigorous steady modality using HYSYS Aspen Version 8.8 considering Khormala crude oil as the specific case of study.
- Determine the most suitable operational conditions to apply in Atmospheric Distillation Unit based on sensitivity parameters using simulation package.
- Propose suitable conditions for improving and maximizing the production of kerosene fraction from Khormala crude oil.

2. OVERVIEW OF DISTILLATION PROCESS AND LITERATURE REVIEW

Atmospheric and vacuum distillation is one of the first steps to oil in its refining. This consists of taking advantage of the physical properties of chemical raw materials for separating them into fractions at different characteristics, without introducing changes in the original chemical structure of components in the atmospheric distillation. Fractions in the atmospheric distillation process are done based on the differences in volatility since this distillation process is performed using different boiling points of the components of crude oil (Wauquier, 2000).

Its application is in the separation of crude oil into intermediate products of specific quality. The obtained products are as follows: not gas condensable, liquefied gases, gasoline, and gasoline for the reformation, kerosene, heavy diesel, diesel and reduced raw. To obtain these products, one or more distillation stages depending on the raw material can be applied. The first stage is usually an atmospheric distillation (low pressure) and the following stage is a distillation at negative pressures (below the atmospheric pressure) which is known as a vacuum distillation.

According to Gomez (2013), the majority of the products obtained in the different stages of distillation are susceptible of reprocess, either for obtaining other fractions by processes of conversion and separation or for improving their quality. Majority of separations is performed by distillation. The use of solvents can isolate a fraction by taking into account relative solubility without regarding their volatility; this type of separation is carried out in several stages depending on the desired degree of separation.

Atmospheric distillation is first refining process which has a raw crude after dehydration, is a key stage and essential in refining; any of a crude distillation fractions thereof, does not generate a compound but groups of compounds, referred to interchangeably as cuts or fractions.

Nuhu et al. (2004) performed a technical investigation of Crude Distillation Unit of N'djamena Refinery Company in Chad Republic. In the referred Crude Refining Unit, the expected work, the lost and shaft work were found to be 2.40E+08, 4.29E+08 and 6.69E+08 Btu/hr, respectively. Likewise, Nuhu et al. (2004) ascertained the second law efficiency to be 35.8 %.

The natural quality of results of a fractionation segment was examined by considering diverse technical conditions of the section utilizing normal gas condensate as a system feeding. The principal configuration was on a conventional refining segment while the following examinations were done altering the refining column section to yield a similar quality of results by keeping the material balance invariable. This investigation incorporates the subtle elements quality variety alongside the variety of plan. This investigation incorporates the elements quality changes alongside respective design changes. The entire study and investigation was done by Rahman and Kirtania (2011) by using Aspen HYSYS 7.1 and a retrofit plan technique and simulation structure used to incorporate unrefined petroleum were carried out by Mamdouh et al. (2013) using HYSYS to simulate refinement of crude oil. The increase of gasoline production in every one of the refineries is the main goal. When focusing on the CDUs is primary objective, optimizing the yield of gasoline and its intermediates will affect positively on total inventory gasoline production. Okeke & Osakwe-Akofe (2009) utilized HYSYS software to develop a simulation of a process and a strategy for the improvement and management systems and operability of each existing associated new production facilities through an integrated atmosphere of various technologies. Such an integrated atmosphere not just creates opportunities for operational deciding but also additionally is coaching tool for the novice engineers. It allows them to use engineering experience to resolve challenges distinctive to the industries in a very safe and virtual atmosphere and additionally assists them to urge inform with the present management systems and to know the basics of the plant operation (Yela, 2009).

2.1 Fundamentals and Principles of Crude Oil Distillation Units

According to Matar (2000), the organic compound intermediates are created by subjecting crude oils to varied process schemes. These embrace a primary distillation step to separate the oil-complicated mixture into less complicated fractions. These fractions are primarily used as fuels.

However, a tiny low proportion of those streams is used as secondary raw materials or intermediates for getting olefins, diolefins and aromatics for petrochemicals production. Additional process of those fractions could also be needed to vary their chemical composition to the desired product.

These new products may additionally be used as fuels of improved qualities or as chemical feedstocks. For instance, reforming a hydrocarbon fraction catalytically produces a reformate wealthy in aromatics. The main use of the reformate is supplement the gasoline pool owing to its high octane number. However, the reformate will not additionally be extract of the aromatics for petrochemicals use. At now, the assembly of intermediates for petrochemicals is not dissociable from the assembly of fuels.

In this section, the assembly of organic compound intermediates is mentioned in conjunction with very different petroleum process schemes mistreatment atmospheric distillation. These embrace physical separation techniques and chemical conversion processes.

In this order of concepts, atmospheric distillation separates the crude oil complicated mixture into entirely different fractions with comparatively slim boiling ranges. In general, separation of a combination into fractions is predicated totally on the distinction within the boiling points of the parts.

According to the optimization (2011), one or additional fractionating columns are used in atmospheric distillation units. Distilling a crude starts by preheating the feed by exchange with the new product streams. The feed is heated more to concerning 320 °C because it passes through the pipe heater (pipe still heater). The hot feed enters the fractionator that commonly contains 30–50 fractionation trays. Steam is introduced at the lowest of the fractionator to strip off lightweight parts (see flow diagram shown in Figure 2.1).

The potency of separation may be a perform of the quantity of theoretical plates of the fractionating tower and the reflux quantitative relation. Reflux is provided by condensation of vapors at the condenser mounted at tower overhead. Reflux quantitative relation is that the quantitative relation of vapors condensation back to the still to vapors condensation out of the still (distillate). The upper the reflux quantitative relation, the higher the separation of the mixture.

Products are withdrawn from the distillation tower as facet streams, whereas the reflux is provided by returning some of the cooled vapors from the tower overhead condenser. Further reflux may be obtained by returning a part of the cold facet stream product to the tower.

In application, the reflux quantitative relation varies over a large ranges depending on the particular separations desired. The uncondensed gases are separated in the overhead condenser; therefore, the condensed lightweight hydrocarbon liquid is withdrawn to storage directly. Heavy hydrocarbon, kerosene, and fuel oil are withdrawn as facet stream products also. Table 2.1 shows the approximate boiling ranges for crude fractions. The residue is separated at the bottom of the distillation tower and will be used as a fuel. It could even be charged to a vacuum distillation unit for chemical changes by cracking or steam cracking process.

2.2 Material Balance of Crude Oil Distillation Units.

According to Perry et al. (1997), a distillation is outlined as an equilibrium-staged separation method within a liquid or vapor mixture or each containing two or a lot of components which are separated into its component fractions of desired purity by the applying and/or removal of warmth.

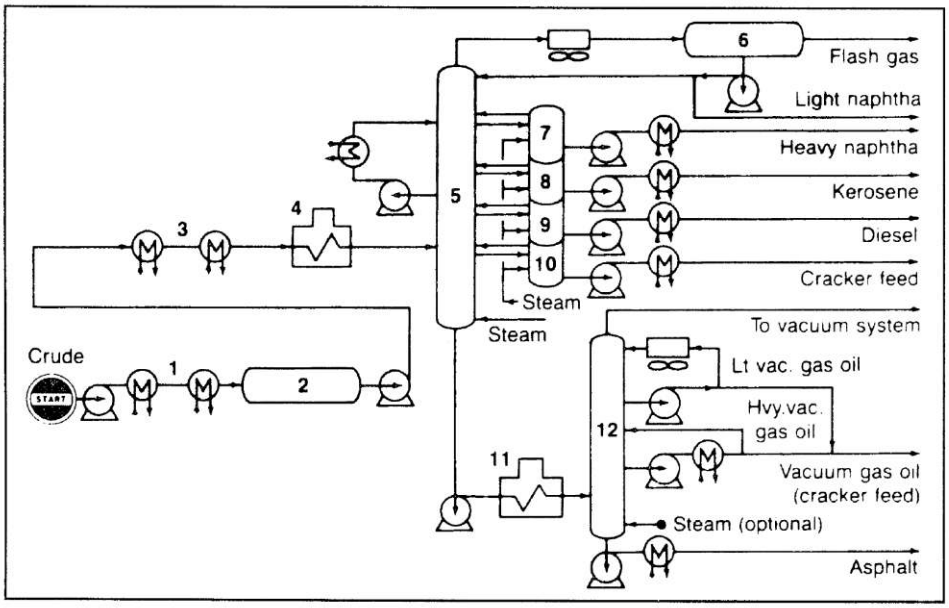
This method is predicated on the equilibrium stage conception, which implies vapor–liquid equilibrium when they are contacted one another. The vapor and liquid streams leave a stage or unit in the saturated form with a lot of volatile part (has lower boiling point) and the less volatile part (has higher boiling point) severally. The separating agent for distillation is heat; therefore, the mechanism of distillation is by the addition of warmth. The necessary variables that have an effect on part equilibrium in distillation are temperature, pressure, and concentration.

Table 2.1 Approximate ASTM boiling point ranges for crude oil atmospheric fractions

Fractions	Boiling Ranges	
	F	˚C
Light Naphtha	85-210	30-99
Heavy Naphtha	190-400	88-204
Kerosine	340-520	171-271
Atmospheric Gas Oil	540-820	278-438
Vacuum Gas Oil	750-1050	399-566
Vacuum Residue	+1000	+538

2.3 Calculation Methods of Distillation

According to Robles (2011), there are two types of distillation columns: staged and packed columns. For staged and packed distillation columns, calculations are of the equilibrium-based method and the rate-based method, respectively.



(1,3) heat exchangers; (2) desalter, (3,4) heater; (5) distillation column, (6) overhead condenser, (7–10) pump around streams, (11) vacuum distillation heater; (12) vacuum Column.

Figure 2.1 Flow diagram of atmospheric and vacuum distillation units

1. Staged columns in distillation

Distillation may be carried out in staged or plate columns in which each plate provides intimate contact between vapor and liquid in continuous countercurrent flow. Each plate constitutes a single stage where there is simultaneous partial condensation of vapor and partial vaporization of liquid.

2. Packed columns in distillation

Distilling columns dumped with packing material is an alternative arrangement to plate columns. The packing material provides high interfacial area for the exchange of the components between the vapor and liquid phases. A true countercurrent flow of vapor and returning liquid (reflux) occurs in packed columns, in contrast to the stage-to-stage flow in plate columns.

Packed columns are used for smaller diameter columns since it is expensive to build a staged column operated properly in small diameters. Packed towers have the advantage of a smaller pressure drop and are, therefore, useful in vacuum fractionation. Another advantage of packed columns is that they can be used to process corrosive materials.

2.3.1 Fractional Distillation

The industrial distillation column is a series of units in which two processes of partial vaporization and partial condensation occur simultaneously. The liquid from a stage flows down to the next stage where it is contacted with the rising vapor. This contact of liquid and vapor flowing counter currently with each other is repeated within the entire distilling column. In analyzing the distilling column in terms of calculation, it may be divided into three sections, as indicated in Fig. 2.2.

Material balances can be made about each of the above three sections of the distilling column, resulting in operating-line equations that relate the concentrations of the vapor and liquid streams passing each other in each stage. An overall material balance at the top section of the column gives the rectifying operating line; balances for the bottom section give the stripping operating line; and balances for the feed stage give the feed line equation. With aid of these operating lines, the calculations for the amounts of streams entering and leaving the column can be made. When McCabe-Thiele method or the Ponchon-Savarit method are used to analyze distillation column, calculations can be graphical. On the other hand, it can be analytical while the stage-to-stage method is used.

2.3.2 Rectifying Operating Line

An overall material balance at the top section of the distilling column results in following equation.

$$V_{n+1} = L_n + D \tag{2.1}$$

A total material balance on the more volatile component is given by

$$y_{n+1}V_{n+1} = x_n L_n + x_D D \tag{2.2}$$

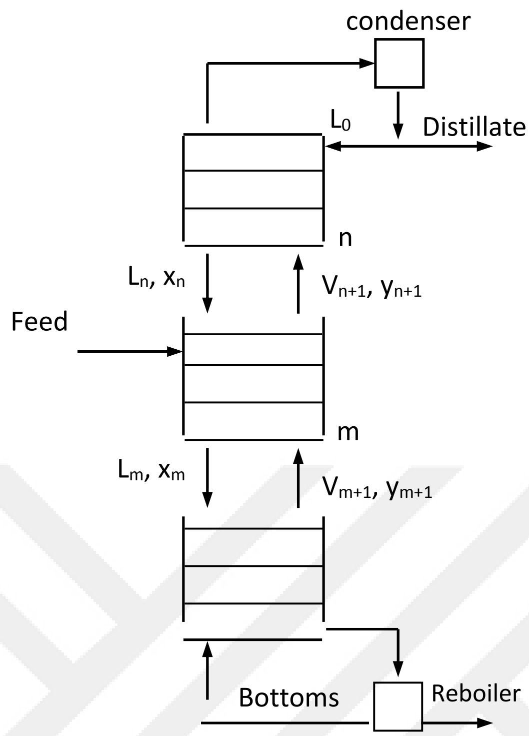


Figure 2.2 Fractional distillation columns

From these two material balances, the following rectifying operating line equation results in

$$y_{n+1} = \frac{L_n}{L_n + D} x_n + \frac{x_D D}{L_n + D} \tag{2.3}$$

With the assumption of a constant molar flow $L_0 = L_1 = L_2$

Equation (2.4) is commonly expressed in terms of the reflux ratio ($R = L_0/D$) as follows:

$$y_{n+1} = \frac{R}{R+1} x_n + \frac{x_D}{R+1} \tag{2.4}$$

2.3.3 Stripping Operating Line

An overall material balance at the bottom section of the distilling column gives

$$V_{m+1} = L_m + B \tag{2.5}$$

A total material balance on the more volatile component is given by

$$y_{m+1}V_{m+1} = x_m L_m + x_B B \tag{2.6}$$

From these two material balances, the following stripping operating line equation results.

$$y_{m+1} = \frac{L_m}{L_m + B} x_m \quad \frac{x_B B}{L_m + B} \tag{2.7}$$

2.3.4 Feed Line

An overall material balance at the feed section of the distilling column gives

$$F + V_{m+1} + L_n = V_{n+1} + L_m \tag{2.8}$$

A total material balance on the more volatile component is given by

$$x_F F + y_{m+1}V_{m+1} + x_n L_n = y_{n+1}V_{n+1} + x_m L_m \tag{2.9}$$

From these two material balances, the following feed line equation results.

$$y_{m+1}V_{m+1} + y_{n+1}V_{n+1} = x_m L_m + x_n L_n + x_F\left(V_{n+1} - V_{m+1} + L_m - L_n\right) \tag{2.10}$$

Graphical representation of these lines is given in Figure 2.3.

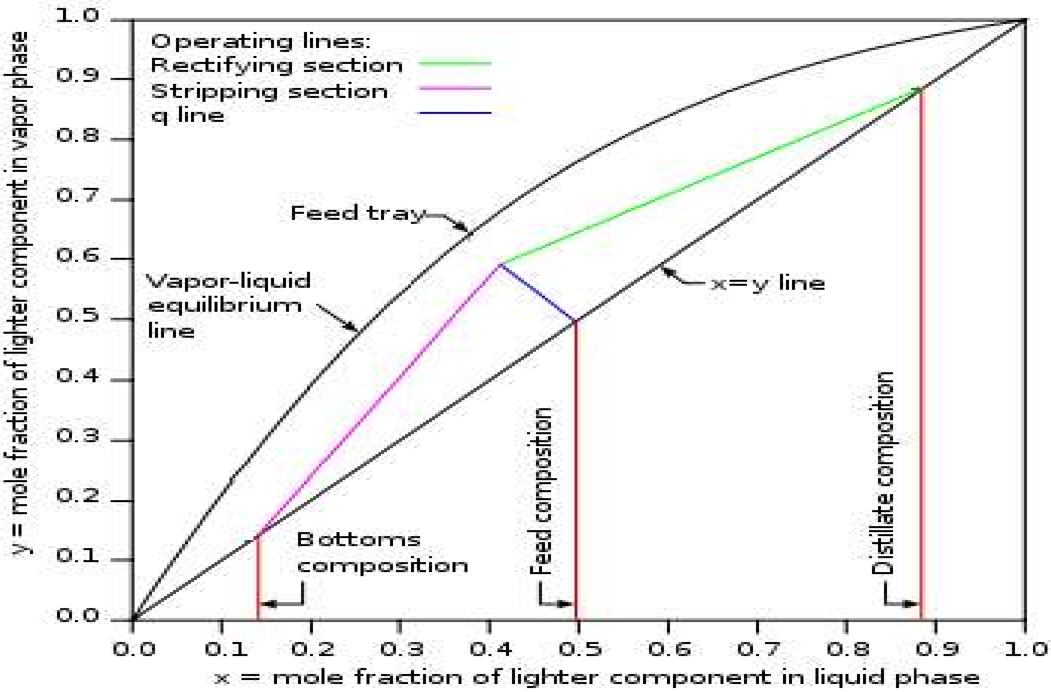


Figure 2.3 McCabe-Thiele method

Another form of equation (2.11), in terms of the liquid content of the feed, q, is given as follows:

$$y = \frac{q}{q-1}x + \frac{x_F}{q-1} \qquad \text{(feed line or q-line)} \tag{2.11}$$

2.3.5 Total Reflux for McCabe-Thiele Method

At total reflux, $D \approx 0$ and so

$$R = \frac{L_0}{D} = \frac{L_0}{\approx 0} = \infty \tag{2. 12}$$

and the y-intercept of the rectifying operating line becomes 0.

$$\frac{x_D}{R+1} = \frac{x_D}{\infty + 1} = 0 \tag{2. 13}$$

Therefore, the overall operating line is the 45° diagonal line. This results in the number of stages N being minimum.

at total reflux ($R = \infty$), ∴ the number of stages $N = N_{\min}$ (2.14)

2.4 Variation of Conditions for Fractionation using McCabe-Thiele Method

2.4.1 Two-Feed Stream Distillation

In distillation with two feeds, the overall operating line includes a third operating line in addition to top and bottom operating lines. This third line is the middle operating line that can be obtained by making material balances around the upper section of the distilling column that includes the first feed stream (positioned above the second feed stream).

2.4.1.1 Distillation with Side Stream

If a product of intermediate composition is required, a side stream may be withdrawn. Three additional variables are necessary such as flow rate, type of side draw (vapor or liquid), and location or composition. The operating line for the middle section can be derived from material balances around the upper or lower section of the distilling column that includes the side stream.

2.4.1.2 Partial Condensers

A partial condenser converts only part of the overhead vapor stream to liquid and returns this liquid as reflux. The remaining vapor is withdrawn as the distillate product. The partial condenser acts as one equilibrium stage.

2.4.1.3 Total Reboilers

A total reboiler converts all the liquid from the bottom of the distilling column to vapor, which, in turn, is returned to the bottom of the column. The material balances and the bottom operating line with a total reboiler are exactly the same as with a partial reboiler. The only difference is that a total reboiler is not an equilibrium stage.

2.5 Ponchon-Savarit Method

If the assumption of constant molar flow of McCabe-Thiele method is no longer valid, the solution to distillation problems is to solve material and energy balances simultaneously on each stage in the distilling column. This solution can be done graphically by Ponchon-Savarit method (Fig. 2.4) in which enthalpy-compositions diagram are used, enthalpy being the vertical axis and composition the horizontal axis.

Figure 2.4 shows the enthalpy-concentration plot for a binary vapor-liquid mixture of A and B at a given constant pressure. The plot is based on arbitrary reference states of liquid and temperature, and it considers latent heats, heats of solution or mixing, and sensible heat of the components of the mixture.

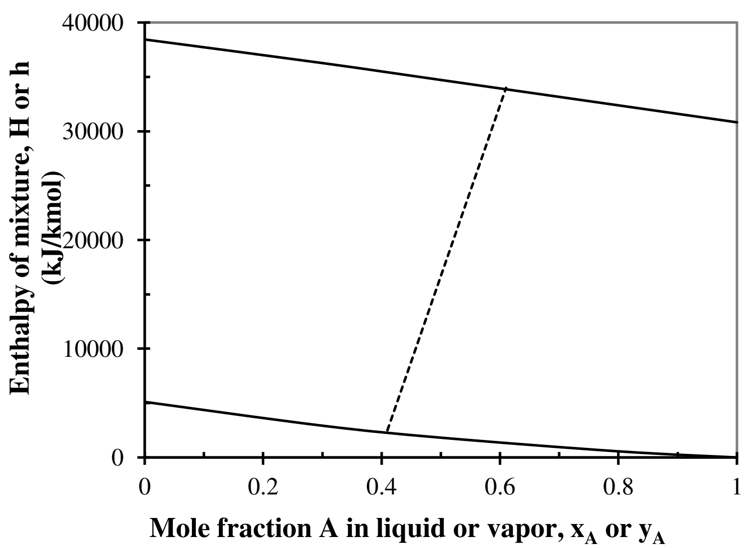


Figure 2.4 Ponchon-Savarit enthalpy-concentration diagram

2.6 Analytical Calculation Methods for Distillation

2.6.1 Total Reflux Ratio

If the relative volatility, α, of a binary mixture is approximately constant, the Fenske equation can be used to calculate the minimum number of stages for case of using a total condenser as explained in the book by Jones and Pujado (2006).

$$N_m = \frac{\log \frac{x_D}{1 - x_D} \frac{1 - x_B}{x_B}}{\log \alpha} \qquad \text{Note: } \alpha = \sqrt{\alpha_{top}\alpha_{bottom}} \tag{2.15}$$

where N_m = minimum number of stages

α_{top} = relative volatility at top temperature

α_{bottom} = relative volatility at bottom temperature

2.6.2 Minimum Reflux Ratio

According to Gavhane (2008), if the relative volatility (α) of a binary mixture could be taken as constant, R_m may be obtained analytically using the Underwood equation for ideal mixtures as follows:

$$R_m = \frac{1}{\alpha - 1} \quad \frac{x_D}{x_F} + \frac{\alpha(1 - x_D)}{1 - x_F} \tag{2.16}$$

Here R_m denotes for minimum reflux ratio.

3. SIMULATION AND PLANT DESIGN FOR DISTILLATION PROCESS

ASPEN HYSYS manual process simulation can be employed for the planning, development, analysis, and improvement of technical processes such as chemical plants and complicated chemical processes, environmental systems, power plants, advanced producing operations, biological processes, and similar technical functions (Agrawal, 2012).

Process simulation could be a model-based illustration of chemical, physical, biological, and alternative technical processes or unit operations in software package. Basic conditions are a radical information of chemical and physical properties such as pure parts and mixtures, reactions and mathematical models permit the calculation of chemical and physical properties used in the method given in simulation program (Robertson and Palazoglu, 2011).

Process simulation software package describes processes in flow diagrams wherever unit operations are positioned and connected by product or edict streams. The software package should solve the mass and energy balance to search out a stability in operation purpose. The goal of a process simulation is to search out optimum conditions for an examined process unit. This can be primarily an improvement drawback, which should be solved in a repetitive process (Denn, 2004, Casavant and Cote, 2004).

Process simulation perpetually use models that introduce approximations and assumptions; but permit the outline of a property over a large variation of temperatures and pressures, which could not be lined up by real information. Models conjointly permit interpolation and extrapolation - among bound limits - and alter exploration for conditions outside, that is, the variation of identified properties.

Process simulation could be a key activity in process engineering covering the full life cycle of a process, from research & development to abstract design and plant operation. During this context, flow sheeting could be a general description of fabric and energy streams in a very process plant by means that of simulation with the scope of planning the plant or understanding its operation.

Steady state flow sheeting is an everyday tool of the chemical engineer. The generalization of the dynamic simulation within the design application is the next challenge. By means that of a capable business flow sheeting system, it is attainable to provide a comprehensive computer image of a running process, a plant simulation model, which might mix each steady state and dynamic simulation. This tool is especially valuable in understanding the operation of a complex plant and on this basis; it will serve for continuous rising the method design, or for developing new processes.

According to Rodriguez (2010), process simulation relies on models. A model ought to mirror the fact at the degree of accuracy needed by application. Having a decent information of the modelling background is mandatory for obtaining reliable results and victimizing the software package effectively. The distinction between productive and unsuccessful computer-aided project ought to be attributed to a poor capability of the user. Process simulation require skilled users to gain advantage from the modelling atmosphere, but not inadequate performance of the machine. That is why a retardant simulation should be fastidiously ready.

Flow sheeting remains dominated by the Sequential-Modular design, however incorporates more and more options of the Equation-Oriented answer mode. A restricted variety of systems offers each steady state and dynamic flow sheeting simulators.

The integration of simulation tools is important to deal with the variability of desires in process engineering. It is fascinating to open the access to simulation technology to a bigger range of model suppliers. This may be accomplished by a cooperative approach between the community of users and of software package producers. The supply of simulation systems on net will boost the utilization of simulation technology in a very international atmosphere.

3.1 Modelling of Process Simulation as Control Tools

Estrada (2013) established the event of models for a more robust illustration of real processes that was the core of the additional development of the simulation software package. Model development was completed on the chemical engineering aspect and so on top of things in engineering. Therefore, the advance of mathematical simulation techniques was developed. Process simulation is so one among the few fields wherever scientists from chemistry, physics, computing, arithmetic, and several other engineering fields work along.

Many efforts are created to develop new and improved models for the calculation of properties. This includes, for instance, the outline of thermo-physical properties such as vapor pressures, viscosities, caloric information, etc. of pure elements and mixtures for various apparatuses like reactors, distillation columns, pumps. In addition, many attempts have been made to outline complicated chemical reactions and mechanics.

Two main distinct sorts of models are distinguished in simulation:

1. Calculations and Balances: Rather easy equations and correlations wherever parameters are fitted to experimental information.
2. Prognosticative Methods: wherever properties are calculable exploitation equations and correlations are commonly most well-liked as a result of they describe the property (almost) precisely. To get reliable parameters it is necessary to possess experimental knowledge that are typically obtained from real information banks or, if no information is publically obtainable, from measurements.

Using prognosticative strategies is so cheaper than experimental work and then knowledge from information banks. Despite this massive advantage, the expected properties are ordinarily solely utilized in early steps of the process development to seek out initial approximate solutions and to exclude wrong pathways because these estimation strategies ordinarily introduce higher errors than correlations obtained from real information.

Walters (1973) indicated that process flow diagrams are often generated by linking modeling software package to simulators and process simulation is additionally inspired the additional development of mathematical models within the fields and Hough simulators offer information about the resolution of complicated issues. The information from simulators such as 'what is happening within a pipe' is often processed with nominal engineering to attain a listing of specific line instrumentation and materials and thus, the precise price of a project.

Because of the automatizing a process of moving from simulation to price, chemical engineers will run more simulation cases for energy reduction, optimization or environmental comes and obtain a price out so quickly with a decent degree of accuracy in terms of what's required physically for a project.

This idea permits management and engineers to see all concepts so that they will realize the best resolution lots quicker, and be sure about the price and come back on investment of a project. Currently everyone seems to be therefore strapped for capital that they have to have a cast iron business case for what they are reaching to do, and the simulation and modeling tools extremely facilitate here.

To assist Engineers' curiosity about this kind of cost accounting, HYSYS as its platform of simulation was established (PDVSA, 2010). The HYSYS links to process simulators and automatically generates process flow diagrams and piping and instrumentation diagrams (P&IDs)

and produces line lists and different two-dimensional information. HYSYS industrial plant is another aspen product applicable here. It permits engineers to model main blocks of apparatus and takes line lists from HYSYS or any P&ID and automatically lay out the plant in 3D. It conjointly works out, but long items of pipe and different connected details ought to be and then, with the push of a button, it prices the plant or project.

Another inventive use for the most recent generation of simulation software package attaches HYSYS to the plant system, which sends data to a model of the plant within the simulator and allows calculation of values that are costly or tough to scale.

Quimitec (2010) mentioned that if somebody links a process simulator to a system, the system itself would see what is an expected calculation from engineering thermodynamic models and choose what is a practical expectation for the behavior of a process, which will tell you the way profitable you are at any given moment.

Michel (2012) indicated that the simulator model would recognize a dangerous situation before operator's intuition, which will lead to faster reactions and spending less time off spec.

In conclusion, the short survey about process simulation gives us a message is "there is almost no discipline in chemical engineering that can afford to ignore the importance of process simulation." It is the inevitable part of chemical engineering and engineering in general. Process simulation is like a flashlight in the hands of a chemical engineer guiding one to obtain the best engineering solution.

For example, a Chemstations' client joined the company's ChemCAD model of a plant that was creating bleaching agent to its system. The system uses the simulation software package to predict purity starting of the plant and to predict an optimized feed rate of steam and water. It conjointly permits operators to control panels for steerage and to work out after they ought to modify the feed rates. They will calculate profitableness instantly supported detector information taken and fed through the simulator model and back to the decision-making program.

Hill expressed that linking the simulator to the system provides advantages since the simulator becomes a software package detector, which gives an expected worth for one thing that was previously troublesome or valuable scale. Perry et al. (1997) mentioned that running the simulator during this vein may allow the simulator to look at a bug become an alarm scenario before an operator might spot it.

3.2 Fundamental and Principles of ASPEN-HYSYS

Advanced Systems for Power Engineering, Inc. (ASPEN, 2017) define HYSYS software in its manual as a tool used by engineers to model any type of process for which there is a continuous flow of materials and energy from one processing unit to the next.

HYSYS has been widely used to simulate processes in chemical and petrochemical industries, refining, oil and gas processing, synthetic fuels and power generation plants. Process models are used at all stages in the life cycle of process plants from process development to design and plant operation.

The operator supplies some data as input to the model and the program itself in the process flow sheet creates some thermo-physical properties related to the supplied data. Output from the model is a complete representation of the performance of the plant including the composition, flow rates, properties of all streams and the performance of the process units. During the process simulation, a HYSYS model could be developed as soon as the conceptual flow sheet of the process becomes available. This model may be updated to obtain more information about the process. Even at very early stages of the project, the model can be used to assess the preliminary economics of the process and to study the effect of technological changes on the economics. (Aspen HYSYS Operations Guide, 2005).

The model can help interpret the pilot plant data and allow investigating process alternatives. Once the decision has been made to build a new plant or to modernize an existing plant, the HYSYS models may be used to study trade-offs, to investigate off design operations and to evaluate the flexibility of the plant to handle different feedstocks. Moreover, simulation studies during process design could avoid costly mistakes before committing to plant hardware (Aspen HYSYS user guide, 2005).

Process engineers may use a simulation model to optimize the design of the process by making a series of case studies to ensure that the plant would work properly under a wide range of operating conditions. For an existing plant, a HYSYS model can serve as a powerful tool for plant and process engineers to improve plant operations in terms of increasing yield and to reducing energy consumption.

The model can be employed to determine changes in the operating conditions needed to accommodate changes in feedstocks, product requirements and environment conditions. The model can guide plant operations to reduce the cost and to improve the productivity of the process.

Finally, the model can be used to study possible modifications for debottlenecking or revamping the plant to incorporate technological advances such as improved catalyst, new solvent or process units. The best way to learn basic HYSYS concept is to solve different simulation problems through the detailed step-by-step-instructions, tutorials, examples and case studies. The HYSYS provides examples of simulator applications. The purpose is to assist teaching the HYSYS to engineers and practitioners. The examples may later serve as templates and can be tailored to represent the specific needs of engineers.

ASPEN was developed in a way to make use of the full power of HYSYS by guiding the trainees through the correct, complete and fast creation of the process flow sheet model. Full screen forms, menus, prompts, help and built-in expert knowledge of simulators guide the reader, step by step through all required and optional inputs, to analyze results and to generate reports and graphs. When physical property parameters and models in the examples are appropriate for the problems considered, the reader should select the most appropriate parameters and models for that specific problem. It is highly recommended to use HYSYS user manuals (Aspen HYSYS Simulation Basis, 2005) for in-depth reference data, theory, examples and tutorials.

4. MATERIAL AND METHOD

To define the scope of this research, it is important to recognize the general procedures to achieve precisely. The objectives can be planned methodology to guide interest indispensable and allows confront and verify theoretical view of the problem.

According Hernandez (2008), it should be as the mode of work to be done because there are many strategies for methodological procedure. This refers to the type of study that will be conducted in order to collect the necessary fundamentals of research. For this reason, the current research focuses as field research referred to improve the performance of Crude Distillation Unit by simulation.

Tamayo, (2007) defined the field research as “a plan or strategy designed to get the information you want, in the same place and time when this occurs”. Therefore, the study is named as field research because the data will be obtained directly from reality namely Erbil Refinery site that belongs to KAR GROUP in Kurdistan, Erbil Region. In addition, it is not experimental study because there were no manipulated variables of study.

This research declares at evaluative level based on facts like the standardized processes in Petroleum Manufacturing (Hurtado, 2010). In this work, the researcher intended to replace a state of actual thing with another desired state of affairs. Segovia (2010) also indicate that evaluation studies aim precise study for accomplishment of event determination at standard conditions. This type of research is associated with the determination of conformity and stages of research instead of establishing an independent and dependent variables. Subsequently sensitivity of study can be done by using virtual parameters taken from real field data.

Since its purpose is to expose the studied event to make a detailed listing of features, the results can be obtained two levels depending on the phenomenon and the purpose of the researchers. It is emphasized that this work is evaluative; therefore, it will be represented in a clear way to evaluate the simulation of crude oil distillation unit by applying specialized Aspen HYSYS software.

4.1 Research Design

This work focuses on a field design study, that allows to study the experimental variables with a minimum number of trials. Since the objective of this research is to get a precise simulation, the indicator or more convenient dependent variable needs to be defined from the operational conditions and parameters are shown in Appendix B.

The process identity elements improve after getting results of the simulation report where the material balance of the distillation unit is a mandatory requirement in the application of the field research where several stages has to be accomplished.

The stages of the observed research described above can be summarized in Figure 4.1.

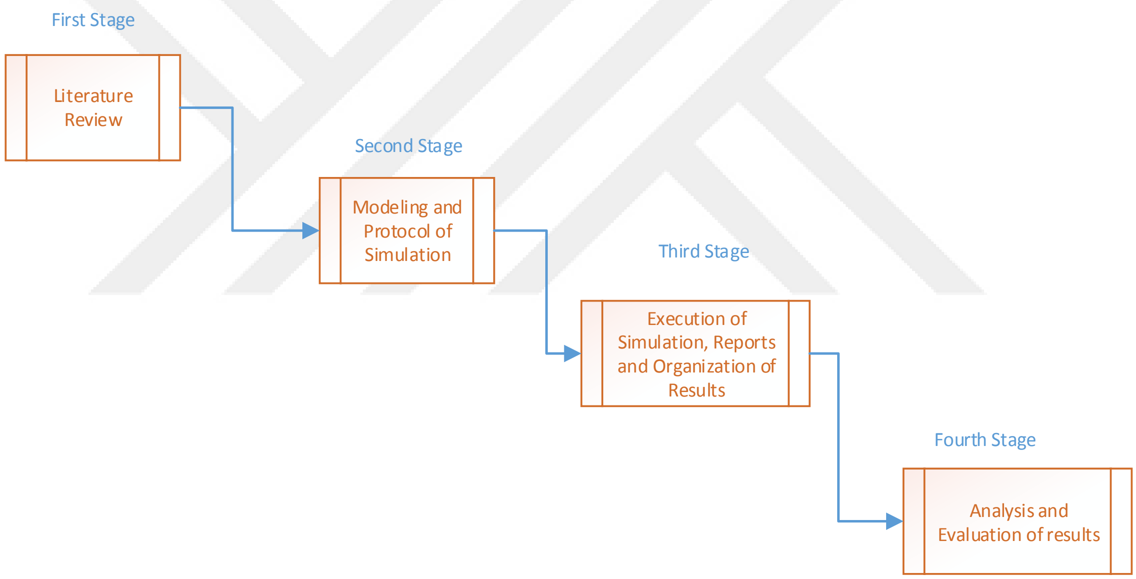


Figure 4.1 Research design for optimization of crude oil distillation unit by simulation

4.1.1 Stage One: Document Reviewed

In this stage, literature is reviewed about the topic under study concerning technical information.

This stage consisted of the following activities:

1. **Research centers, libraries, internet sources visited:** The following sources were consulted to obtain adequate information about topic. a) Egyptian and Americans Journals

of Petroleum, b) International journal of Industrial Chemistry, c) papers from Department of chemical Engineering College at different universities.

2. **Selection of information based on the level of relevance:** Based on fact that research selected was identified by considering next important information. a) Manuals of Simulations over this study, b) procedures for mathematical approaches and ways to apply to crude oil distillation unit.
3. **Specialist personal interview and scientific articles and referent books:** Mechanism of procedures for modeling, simulations, and reports were used to learn how to develop simulation and give guidance to understand opportunities for improving crude oil distillation complex.
4. **Finally, methods for evaluating the effectiveness** associated with evaluation of process itself, this activity gave the theoretical orientation to define the range of optimizable parameters by sensitive studies.

4.1.2 Stage Two: Modeling and Simulation of Khormala Crude Oil

The model for prediction of properties, equilibria and thermodynamically protocol and the procedure for simulation were made in this section. The detailed procedure was done using customized procedure recommended by ASPEN company and the book by Vega (2010). The steps for this procedure are: a) Define thermodynamically model for prediction, b) Unit set, c) Components present in streams, d) Energy stream and, e) Simulation environment.

4.1.2.1 Preparing of Modelling

In order to start any modeling, you have to create the model where the modeling process contains a sequence of events.

4.1.2.1.1 Creation of a New Case

The first thing to do is to open Aspen HYSYS to create a new case and window shown in Figure 4.2.

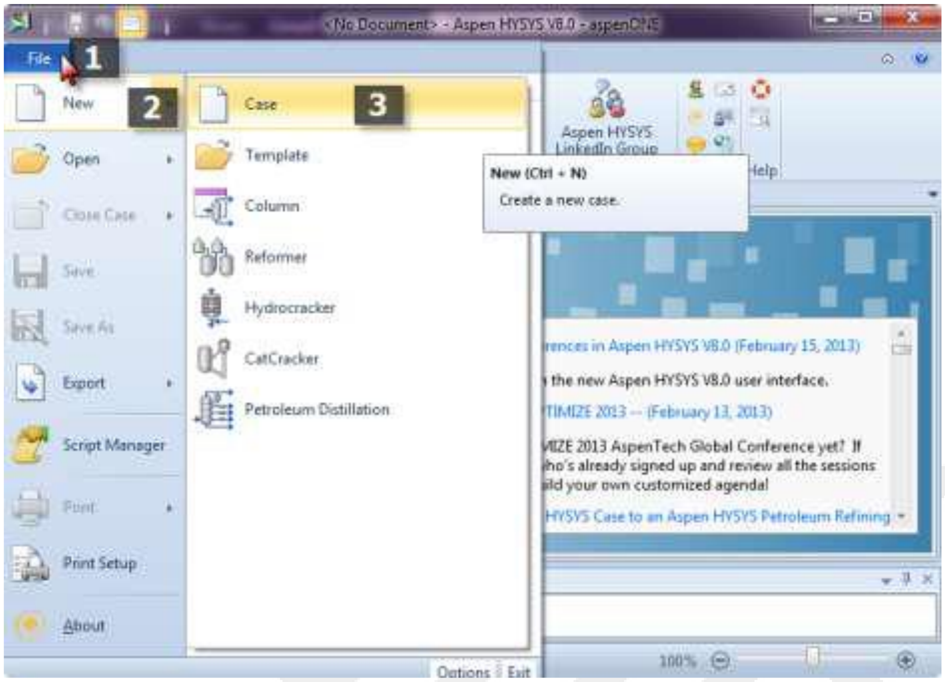


Figure 4.2 Simulation basis manager window

Figure 4.3 is composed of a series of tabs in which these components will be introduced since they will be used to determine characterizations of crude oil examining in the simulation and thermodynamic method in HYSYS.

Other tabs within this window are those hypothetical which serves to create pseudo components or hypothetical compounds and the Reactions tab is used to define the reactions that occur in the simulation. In this case none of these tabs will be used since there is no reaction in the simulation although hypothetical or pseudo compounds are needed. Hypothetical or pseudo compounds will be used to determine crude oil characterization.

Before starting to enter any information, it is important to see the HYSYS preferences menu from the Tools menu.

Same as the previous window, the most of the windows that appear in the HYSYS show several tabs and each tab is composed of a series of pages. Among them, Options and Dynamics within the Simulation tab and the Units page within the Variables tab are highlighted.

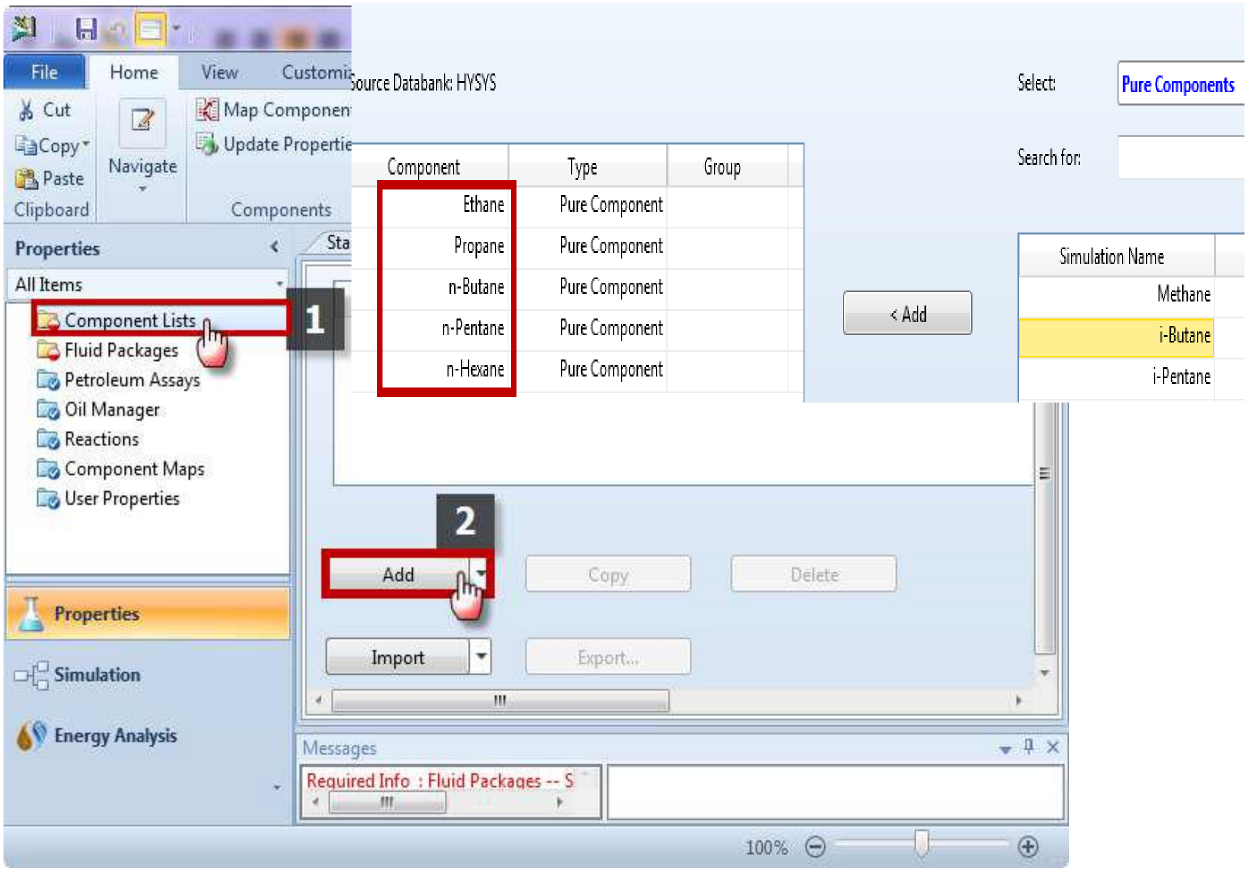


Figure 4.3 Choosing system components from data bank

The Options page serves us mainly to select the general options that are required to activate HYSYS for taking into account the Use Input Experts option that will be of great help for defining the column.

The dynamics page will be used in the dynamic simulation. The Units page helps to select the set of units needed to use. The database of the program contains three Unit Systems, the most of which are unmodifiable, but units of some properties can be changed by copying one of the set of units and then modified the units that interest us. In present case, the Units System Field are copied and modified to change units of some variables.

4.1.2.1.2 Introduction of Components and Choice of Thermodynamic Model

The next step in the simulation is to add the components. The compounds added initially will be the non-oily compound such water and the light ones such ethane, propane, isobutane, n-butane, isopentane and n-pentane.

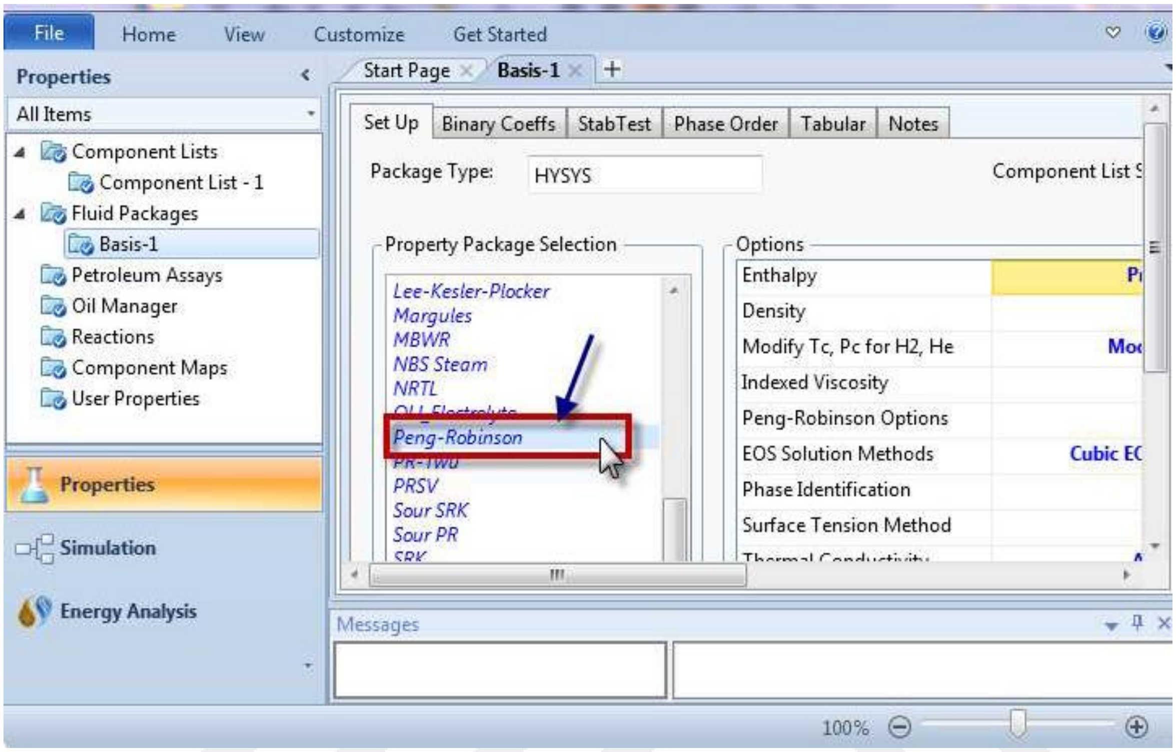


Figure 4.4 Fluid Package selection window, Peng-Robinson selected

Once these initial components are introduced, the thermodynamic method needs to be chosen to incorporate it in the Fluid Packages tab as shown in Figure 4.4. The thermodynamic model is a fundamental part of the simulation with the definition of crude oil.

4.1.2.1.3 Definition of crude oil for simulation

Once the light components and the fluid package are created, the crude oil should be introduced to proceed the simulation. In crude refining dealing with fictitious or pseudo components that are not known or defined compounds whose properties are determined by characterization procedures.

Crude oil is a homogeneous mixture of organic compounds. If their properties and characteristics are very similar, it is very difficult to completely separate from one another, especially in the case of similar boiling temperature range. Due to this and the multitude of different compounds existing in the crude, pseudo components must be created, which will be no more than a grouping of compounds at a certain boiling temperature range.

HYSYS creates these compounds with only the introduction of a series of data easily obtainable by laboratory tests of a crude sample. In present case, there is a TBP distillation curve after giving data that HYSYS needs to create and characterize the crude oil. The necessary data needed for HYSYS to create crude oil is given in Appendix B.

To transform these data into the characterization of crude in the HYSYS, there is Oil Manager tab as shown in Figure 4.5. The thermodynamic model must be selected before creating pseudo components. For this reason, the fluid package has defined previously.

Next step is introducing the data obtained from the laboratory and then HYSYS creates the curves by using the supplied laboratory data. The first thing is to define the extrapolation method known as Lagrange method to compute points for all curves. The extrapolation method will be defined in the Calculation Defaults tab. Once this is done, the composition of the light ones and the data extracted from the laboratory test in the Assay tab must be entered.

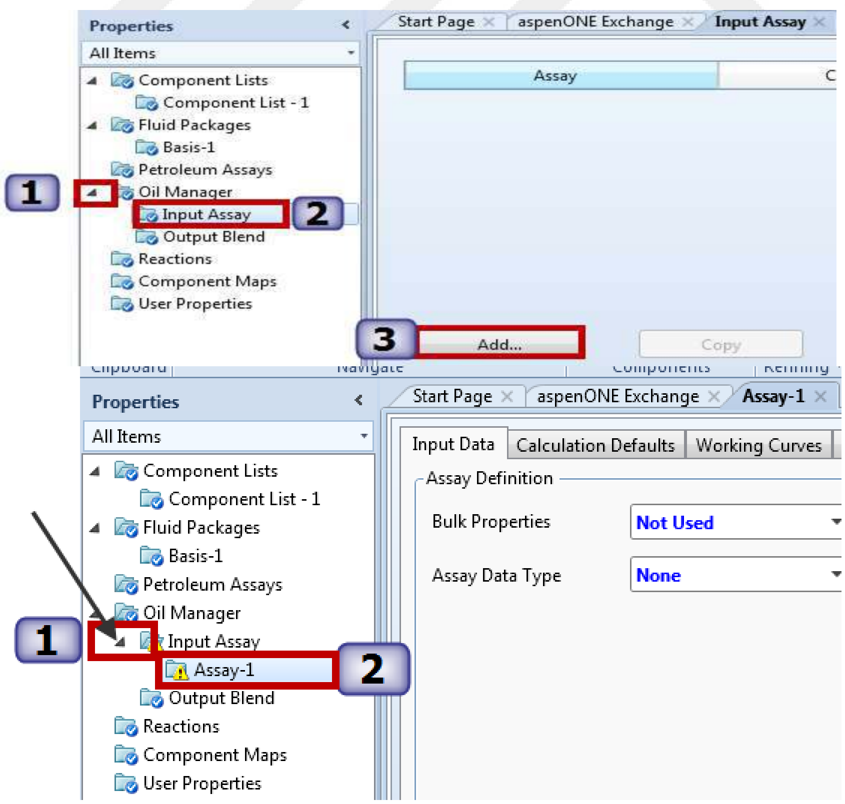


Figure 4.5 Window for the characterization of crude oil

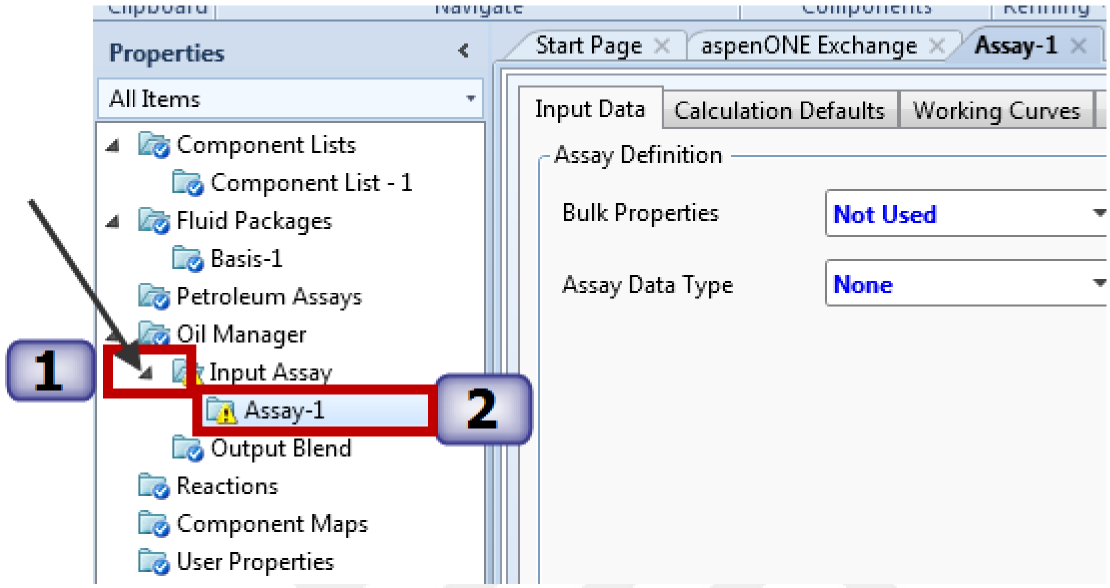


Figure 4.6 Window for the characterization of crude oil

The data introduced to the simulation were given in Appendix B.

To create the used oil, the window appeared in the Figure 4.6 will be used. Most of the tabs in this window are useful for our purpose, either for the input of the data (Input Data and Calculation Defaults) or for the visualization of the results.

4.1.2.2 Simulation Environment

Upon entering the simulation environment shown in the Figure 4.7, the simulation plane appears to show the current created in the installation of the crude oil, which is the one that will supply the information about the power supply to the system. HYSYS is a visually intuitive program and shows different color depending on whether the current is completely defined or not. In this case, it has not been defined yet and appears in light blue but when its defining is finished, it will change to be blue.

4.1.2.2.1 Creation of the Feeding Currents

The first step in the simulation is to create the supply currents to the system and define the conditions and composition.

There are two main ways of introducing the currents as well as the operation units; one is selecting it from the object palette and another through the Workbook. To install a single stream, it is more convenient to do it from the palette. If there are several streams to create, the Workbook is the best option since all currents can be created and defined with their composition a single window. Apart from the composition, the currents need temperature, pressure and flow data for their complete definition as shown in Figure 4.8.

4.1.2.2.2 Definition of the Heating Train Equipment

Once the power currents are created, the next step is to start with the installation of the equipment.

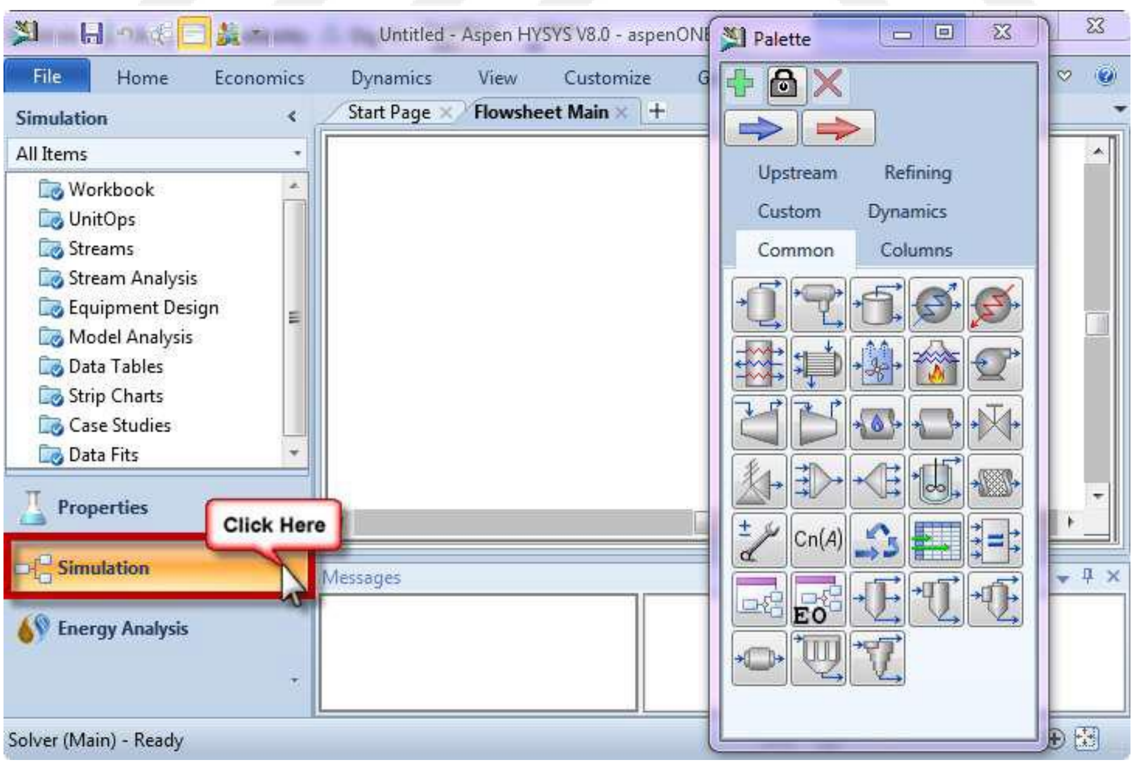


Figure 4.7 Initial image of the simulation environment

As in currents, HYSYS gives different colors to the equipment depending on whether it is completely defined or not.

If the equipment is not defined or does not have the currents connected, it will appear in red. If the equipment is defined but has not simulated, it will appear in gray with a yellow border and if the equipment is correctly defined and simulated, it will appear in colored gray with black borders. The colors of the options can be changed in the preferences menu.

The next equipment that must be installed is the heater; its function is the thermal conditioning of the crude to be reached a certain temperature. When the gaseous current of the flash and the one coming out of the furnace are mixed, it will be reached to the desired temperature for feeding the tower. The type of furnace installed needs the specification of the input current and the output and heat currents (these last two created from the design window of the furnace) as shown in Figure 4.9.

Like the previous equipment, the furnace needs two design specifications; the load loss and the heat provided by the energy current (unknown). This last specification can be calculated by HYSYS program.

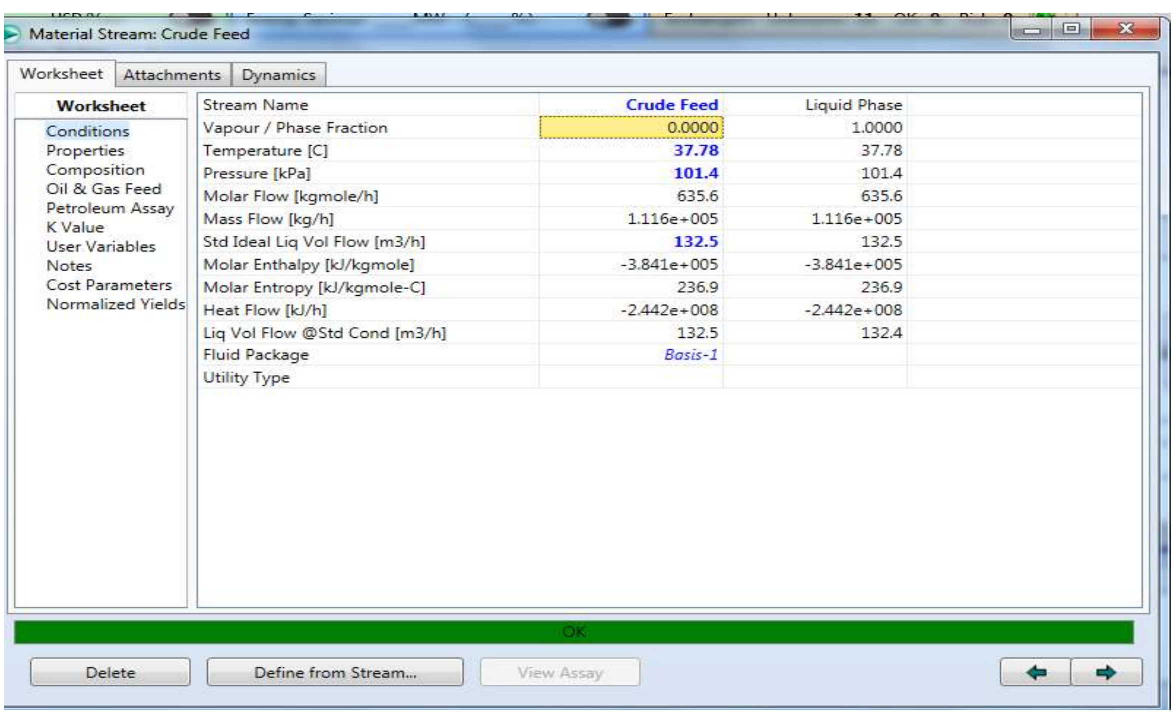


Figure 4.8 Window of the feed stream

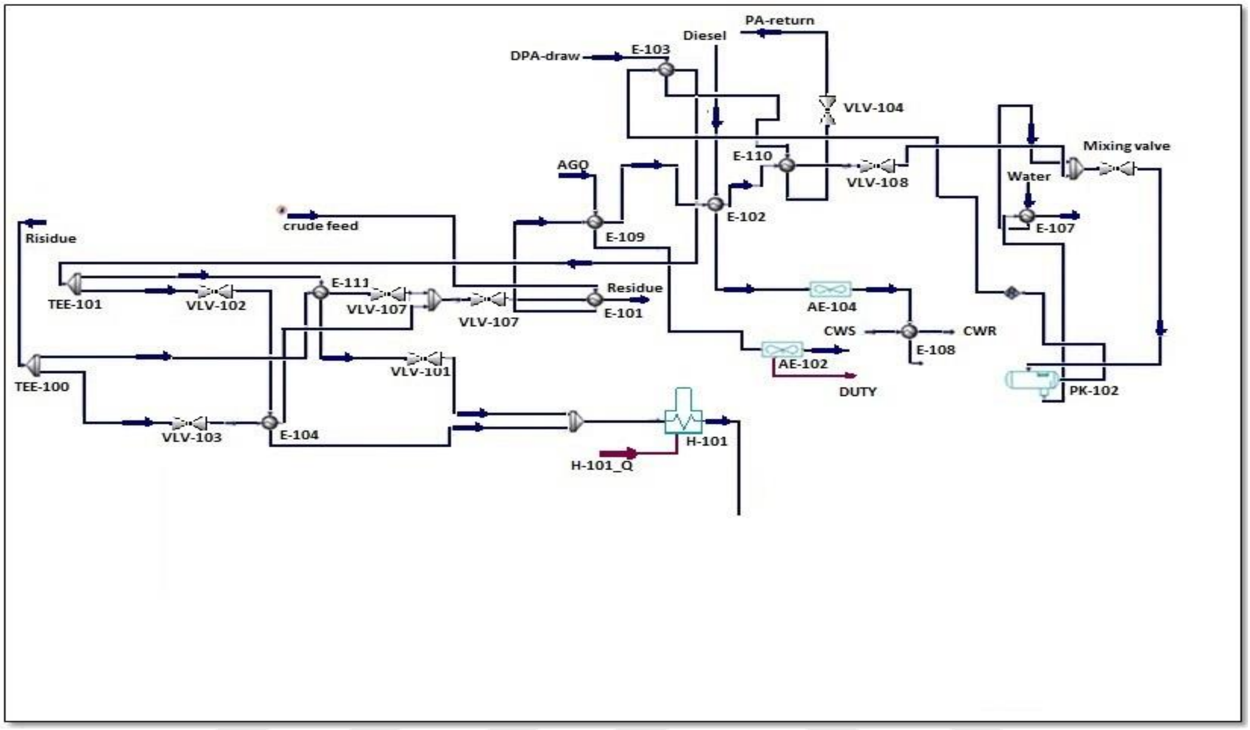


Figure 4.9 Furnace design window

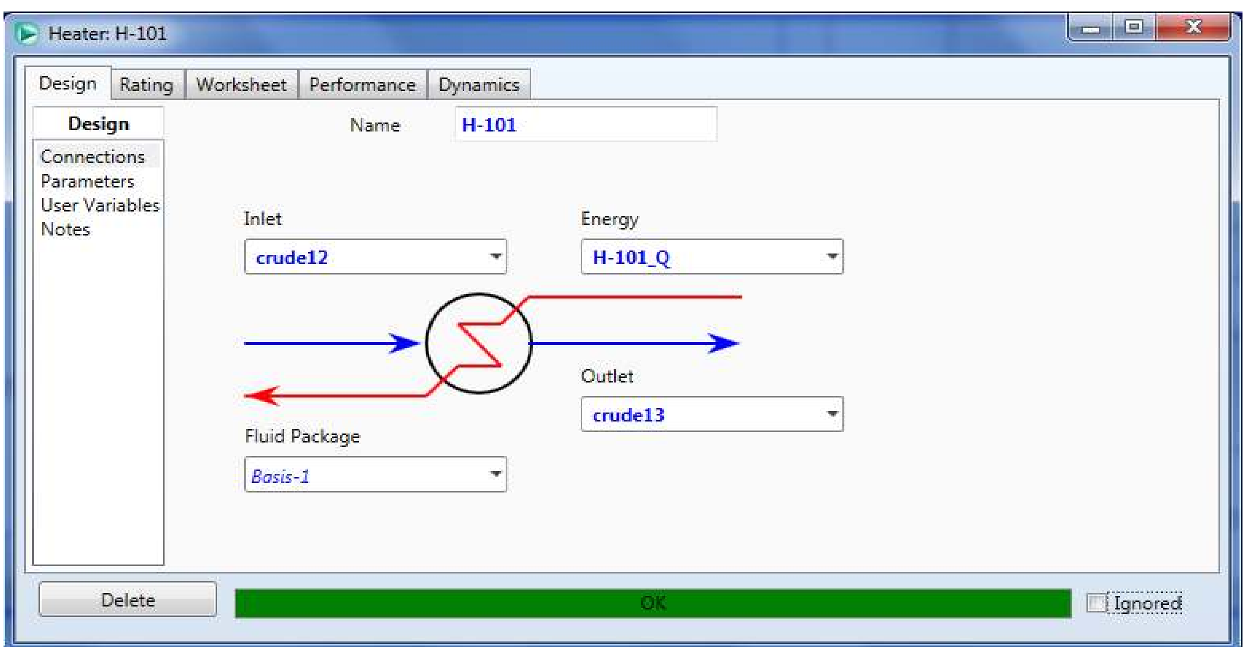


Figure 4.10 Flow diagram until the heater installation

4.1.2.2.3 Installation of the Fractionation Column

After installation of entrained preheaters as shown in Figure 4.10, the most important part of the simulation is the installation of the crude fractionation tower. First, a current of energy that is created will enter the tower along with the crude oil supply. The function of an energy current is to provide the necessary heat to create enough reflux at the feed point and on the plates below it as shown in Figure 4.12. The necessary heat that this current must provide will be calculated with the design of the column.

HYSYS has a number of predefined columns that can be installed and defined completely. The column chosen is a Refluxed Absorber. After that column is installed by using the option Use Input Expert HYSYS guides in the specification of this column in a kind of 4 windows tutorial, the next window will be accessible only if the previous window is correctly completed.

In the first window, the following must be defined:

- Number of theoretical plates: 39.
- Power currents (Power Tower and Q-Power) and input plate (36).
- Steam entering through tail (Steam Bottom).
- Condenser type (Partial).
- Create the product streams (Off Gas, Naphtha, Water and Waste).

The chosen condenser will be partial although it will not have gaseous outlet, eventually it will have two liquid currents: one of them is water and another will contain the naphtha and the final light (Technology, 2004).

The first window allow us to start the numbering of the trays wherever is required, above or below.

The estimated values of condenser output pressure, tail output and pressure loss in the condenser have to be defined on the second window. With these data, HYSYS calculates the distribution of pressures along the column.

The third window asks operator to estimate the temperatures in the condenser in which the upper plate and the last plate temperatures. Since these data are optional, they would not be strictly necessary to specify them.

Concluding these four steps correctly, the necessary specifications which can be found in Appendix B are introduced for the design of the column. The windows are shown in Figure 4.11.

4.1.2.2.4 Installation of Side Stripper

In the installation of the side strippers, the number of plates of the stripper must be specified. Regarding the number of trays, they bibliographically vary between three and six plates, the number chosen for the present case will be three plates in all the strippers as shown in Figure 4.13. For the specification of the flow rates of the product streams, the operator must look at the data obtained by the HYSYS in the creation of the crude and the obtained percentages should be multiplied by the feed flow. Figure 4.13 shows the summary that collects the Side Strippers page of these three installed equipment.

4.1.2.2.5 Installation of the Pump Around

When a lateral extraction is taken place, the stream is passed through a stripper and part of it will be returned to the tower. The temperature of the return stream is higher than the temperature of the mixture in the return plate, which implies a lower liquid phase and influences in the quantity of reflux that may be less than necessary for the best operation of the tower.

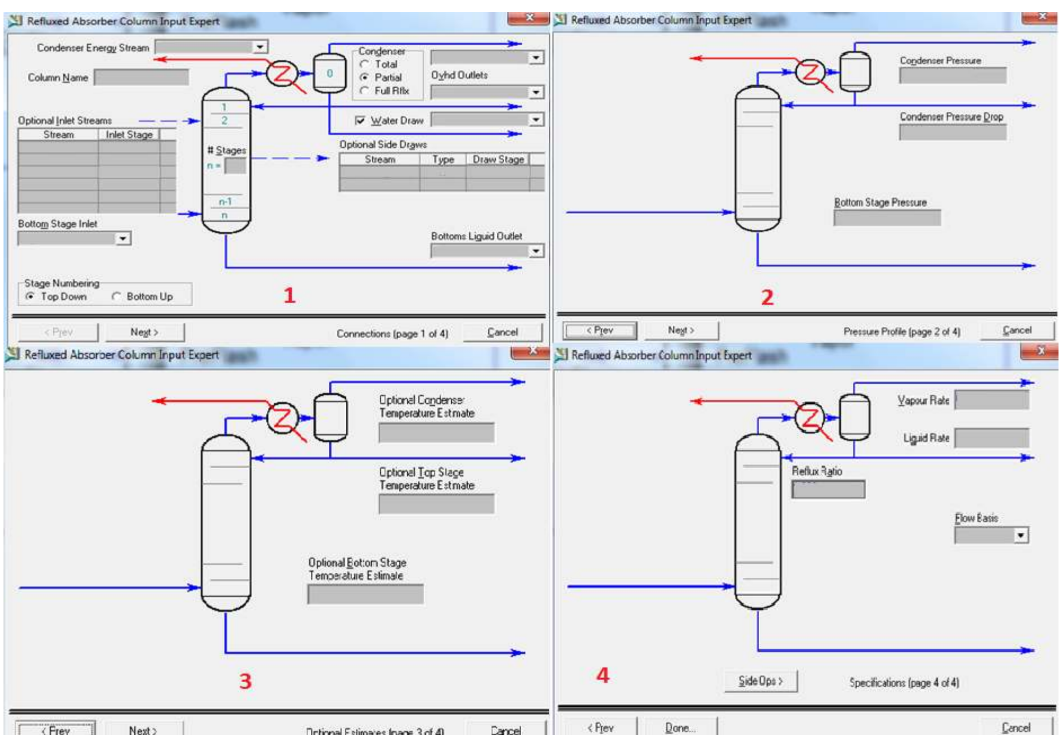


Figure 4.11 Windows to install the column

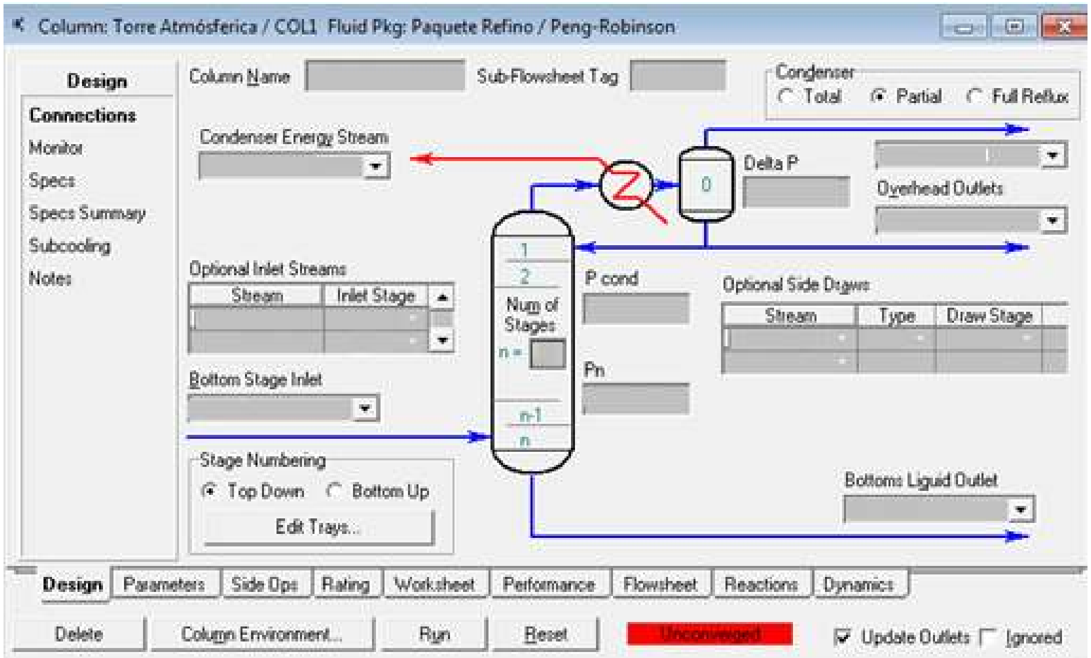


Figure 4.12 Column design window

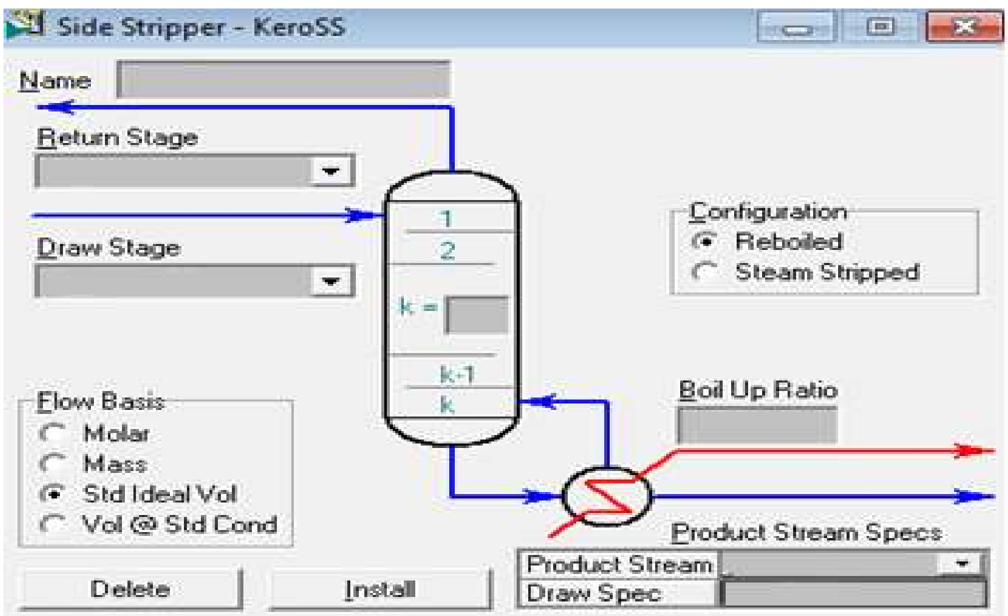


Figure 4.13 Installation window of a side stripper

To correct this, it is necessary to include pump around for lateral extraction for guaranteeing the reflux in all plates of the column. It consists of extracting part of the flow passing through the plate and a heat exchanger and returning to one plate higher at a lower temperature.

When a pump is installed around, HYSYS will require the extraction and return plate and will allow the possibility of introducing a loss of load in the exchanger. Pump around generates two degrees of freedom over the system; therefore, two specifications must be entered. HYSYS by default asks for the flow and the temperature difference as shown in Figures 4.14 and 4.15.

4.1.2.3 Steady State Simulation and Analysis

After finishing as illustrated in Figure 4.16, the obtained results from the static simulation should be checked whether it is acceptable or should be improved. To improve the results it must be known what variables can be modified and how they influence the system. This can be done by choosing case study tool as shown in Figure 4.17.

The flows of the products have been specified to maximize production by adjusting some variables such as temperature and pressure in the distillation column.

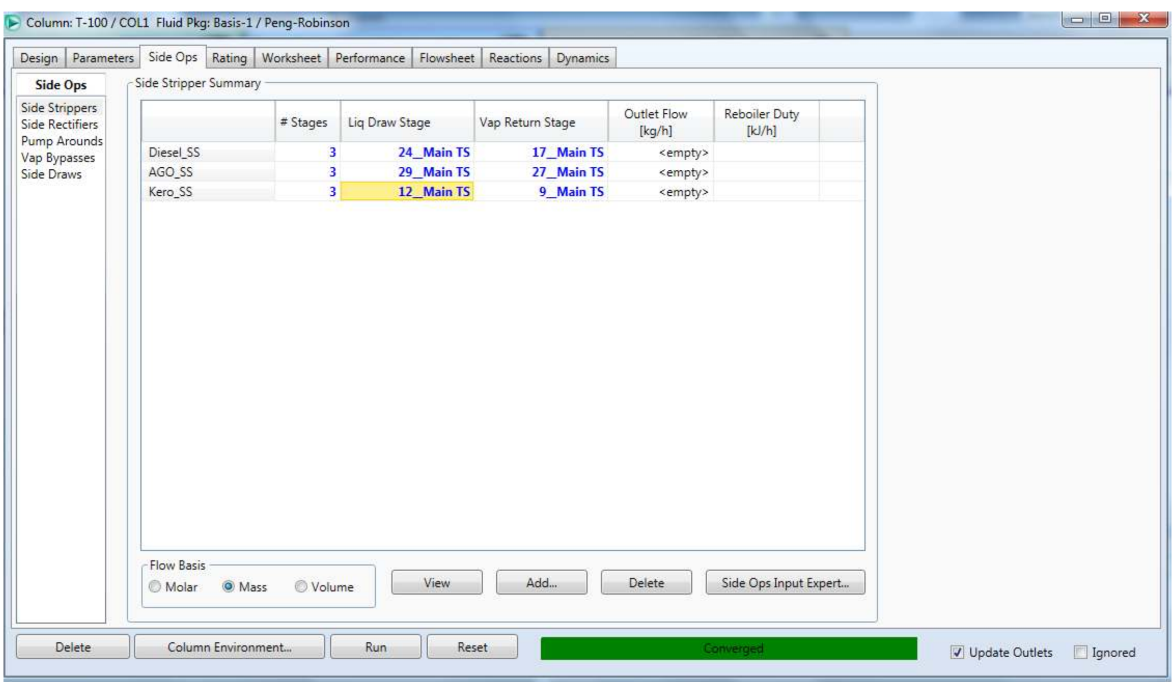


Figure 4.14 Summary table of the lateral strippers

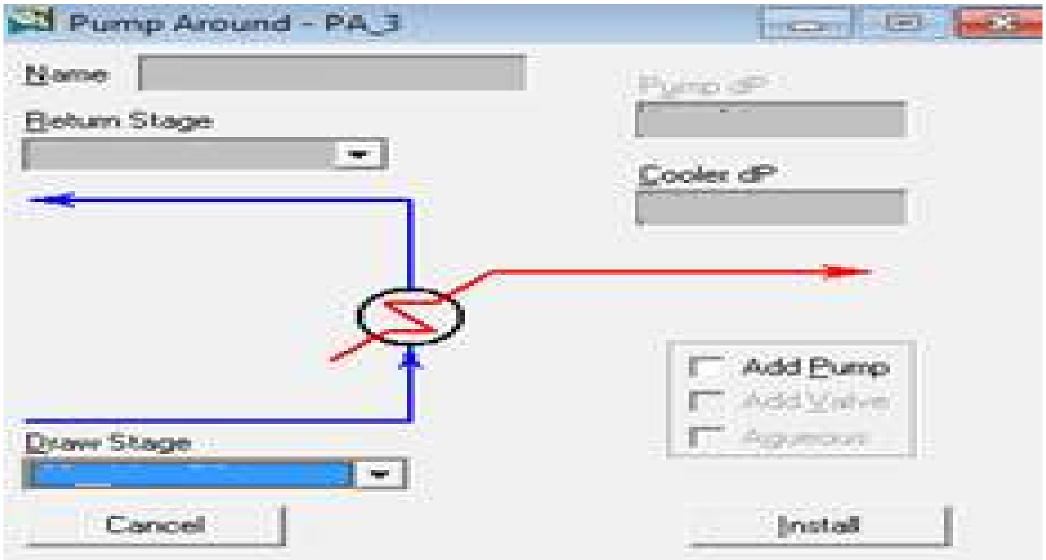


Figure 4.15 Installation window of the pump around

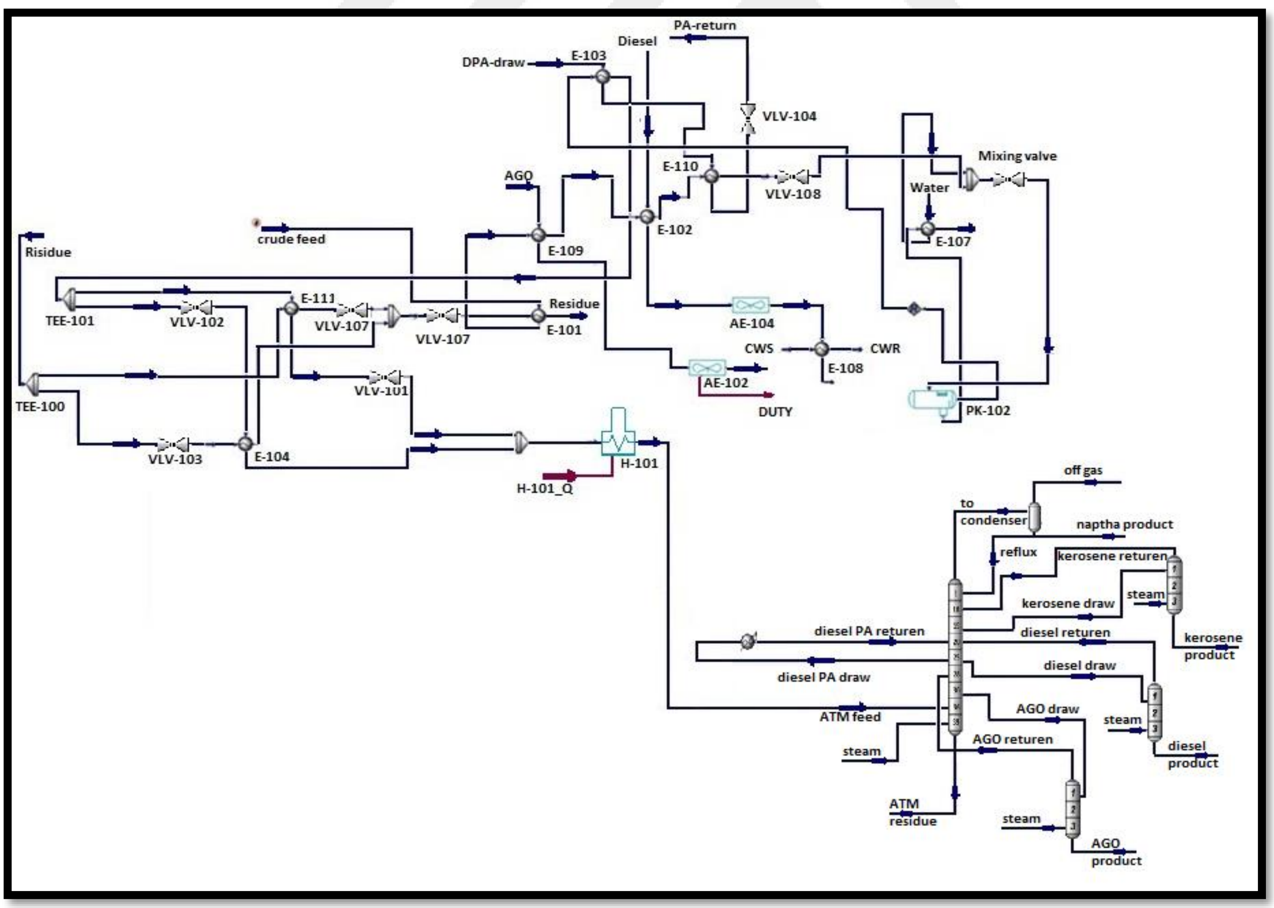


Figure 4.16 Process Simulation model of crude oil distillation section

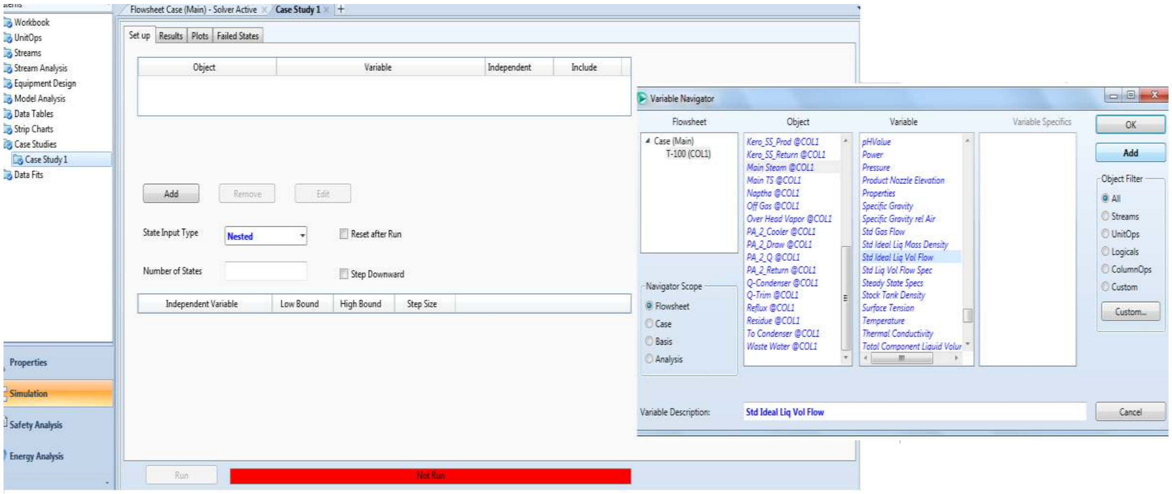


Figure 4.17 Case study window

Other variables must be within a limit to obtain a good product and finally there are the variables that affect the energy saving of the plant such as the consumption of steam. All these variables will depend on a greater or lesser extent on the other variables of the system. The case study helps to see the evolution of the system both graphically and tabular and observe whether the variation of one or more variables produces a positive or negative effect on the others.

Once the variables such as steam flow fate have been created, a scenario must be created to select the parameters that required for the created scenario. When the first state is recorded as the current state and then the value of this variable is changed, the different states will be recorded. Finally, it can be seen that how the variables defined in a table or graph have evolved and reached a consensus on the best value for the manipulated variable.

4.1.3 Stage Three: Report and Organization of Simulation Results

After the simulation runs without errors at total bold process lines. Reports will be generated as a screen reports for simulation. Due to the fact that the used ASPEN license is only for academic purpose, printer report module is not available (only available for commercial company's versions). All results would be presented in different screens among all Chapter of Results and Discussion.

4.1.4 Stage Four: Analysis and Evaluation of Results

At this stage, the results of simulation at normal and steady conditions of Crude Oil Distillation Unit were analyzed by considering the results reported by last step. Those results compared with KAR Quality Specification in order to determine its conformity with international quality control requirements. The validation was conducted by professional KAR and the results of present investigation are presented in Results and Discussion Section.

Secondly, the efficiency of simulation process was measured by material balance comparing actual data with the data obtained from the simulation. The equations for material balances and process diagram of distillation unit were shown in previous figure. Those sensitivity parameters are:

- Heaters operation ranges.
- Stripping steam rates.
- Optimal cutoff temperature for kerosene at distillation tower and its increased production.

4.2 Equipments and Materials

In this section, list of equipments is presented and miscellaneous materials used in simulation package to develop optimization of Distillation Tower are also listed.

4.2.1 Software Package

The Aspen HYSYS (V8.8 (License. HYAC9322456)) simulation program was used to simulate distillation column and investigate various operating parameters. Since HYSYS offers a high degree of flexibility combined with a logical approach, it is an extremely adaptable process simulation tool (Technology, 2005b).

The list of equipment used in this study is shown in the Table 4.1. Technical details of some of them cannot be given because of Refinery confidentiality policy.

Table 4.1 List of main equipments used in this research (own, 2017)

Tools	Uses
HYSYS – ASPEN V8.8	License. HYAC9322456
Process Flow Diagrams	Confidential Information
Khormala Crude Oil Stabilizer.	Trains 1 and 2
Log Process Data	Erbil Refinery (confidential)

4.2.2 Procedures Associated in This Study

The detailed procedure followed in HYSYS V8.8 academic versions is important to mention that some steps were customized for fitting by experience and the process data were taken from operational log data.

Selection of Process Data: In this step, a stable group of data was taken from Operators Log Data coming from daily shifts for one month of normal operations and extreme data or outliners were excluded. Table of results can be seen at Results and Discussion Chapter.

Selection of Process Overview: From PFD (confidential information given by KAR GROUP only for this purpose) was studied, in this part it can be noted that Crude Oil Distillation Unit is suitable to improve. The installed capacity, configuration, process design and constructions details are given in Appendix A.

Definition of Simulation Basis: Based on this specific case, extremely volatile components (C1, C2, C3 and up to C9) among a medium-heavy crude oil like Khormala Crude oil (API range: 33 – 36) were selected. The thermodynamically model, the most suitable for this study, is PENG-ROBINSON Equation of State than Multicomponent equilibria based on VLE.

Definition of Unit Set: In this study were selected two types of units, a) metric value because this makes easier comparison with design and construction references and b) British unit due to the fact that system of units is used to control Distillation Unit by shifts.

Definition of Material Streams: In this step, all material streams from feed, intermediate streams and outlet were defined in terms of temperature, pressure, molar flow, and typical composition. It is important to mention that every stage of separation can simulate the total tower having a preheater.

Definition of Energy Equipments: In order to equalize simulation, the real PFD was necessary to add a heater with different thermal duty and with zero pressure drop (this equipment was defined in that way because tower has all pressure drop of system).

Definition of Material Equipments: In this part, eight preheaters that were added work in series with energy equipments and with total alignment with heaters. This equipment will be working as drums having free condensation of light condensable phase.

Definition of Oil Characterization: Includes lights ends, last Crude Oil Assay available provides TBP curve data.

Definition of Mixer Point: This is a virtual option of simulation that simulates the mixing point of all intermediate streams to get only one gas product outlet.

Definition of Stream Specifications Prior Evaluation: the general design document specification were determined if simulation converged onto real and expected results. In this section, the most important parameters to verify are the petroleum product yields, density, gravity and cutoff range.

Definition and Runout of Side Strippers: Types to be Used, Temperature and Flow

Definition and Run out of Pumps Around: Flows, duty, and effective heat transfers between trays will be needed:

Definition and Run out of Atmospheric Tower: The information supplied to HYSYS is 39 trays and reboiler, rate of condensers, conditions, definition of cutoff distribution.

Definition of Side strippers: AGO, kerosene and diesel side strippers will be added, those equipments will need flow, temperature, and duty. Pumps around will need to work effectively with side strippers to adjust optimization of flow rates and duties.

Sensitivity Studies: In this section, pressures, flows, duties and temperature will be modified in order to achieve maximum yield of kerosene cutoff stream without affecting significantly rest of streams.

5. RESULTS AND DISCUSION

In this Chapter, the results of this research based on goals obtained by specific objectives are presented as follows:

- Modeling and application of simulation.
- Results related to study of sensitivity parameters over distillation unit.
- Evaluation of result's quality by comparison with international specification for processing or commercial purpose of streams produced.

5.1 Crude Oil Characterization

The crude oil was characterized using the true boiling point (TBP) analysis methodology. The analysis was meted out by HYSYS program and Erbil refinery.

5.1.1 Crude Oil Properties

The property analysis of the crude for some months is summarized in Table 5.1. As shown in Table 5.1 there is a variation in properties of crude oil, any tiny variations in composition will greatly have an effect on the physical properties and processing needed for manufacturing marketable products.

Table 5.1 Properties of Khormala blend crude

Test Item	Months						
	1	2	3	4	5	6	7
Density, kg/m^3	856.9	856.3	844	852.2	852.9	845.2	847.8
API°	33.6	33.8	36.2	34.5	34.5	33.1	33.4
Water content, wt %	0.09	0.74	0.3	0.25	0.05	0.08	0.07
Salt content, mg NaCl/L	5	6.2	8.3	9.3	10	8	4.8

5.1.2 Volume Percentage Yield of Crude Oil Products

A summary of the product volume percentage yields from the crude reports for some months is represented in Figure 5.1. The differences in the volume percentage can be attributed to the differences in the feed characteristics.

According to the monthly reports of production, percentage yield of the product volume has been represented in Figure 5.1. It can be seen that the study carried out for Khormala crude oil, which has an API grade of 34, shows a product yield quite close to the theoretical yields for the crude oil of the same API.

As can be seen in Figure 5.1 the percentage yields for the naphtha, kerosene, diesel, AGO and the residue mixed with the cut of light gasoil and used as fuel oil are 27 %, 4 %, 14 %, 3 % and 52 % , respectively.

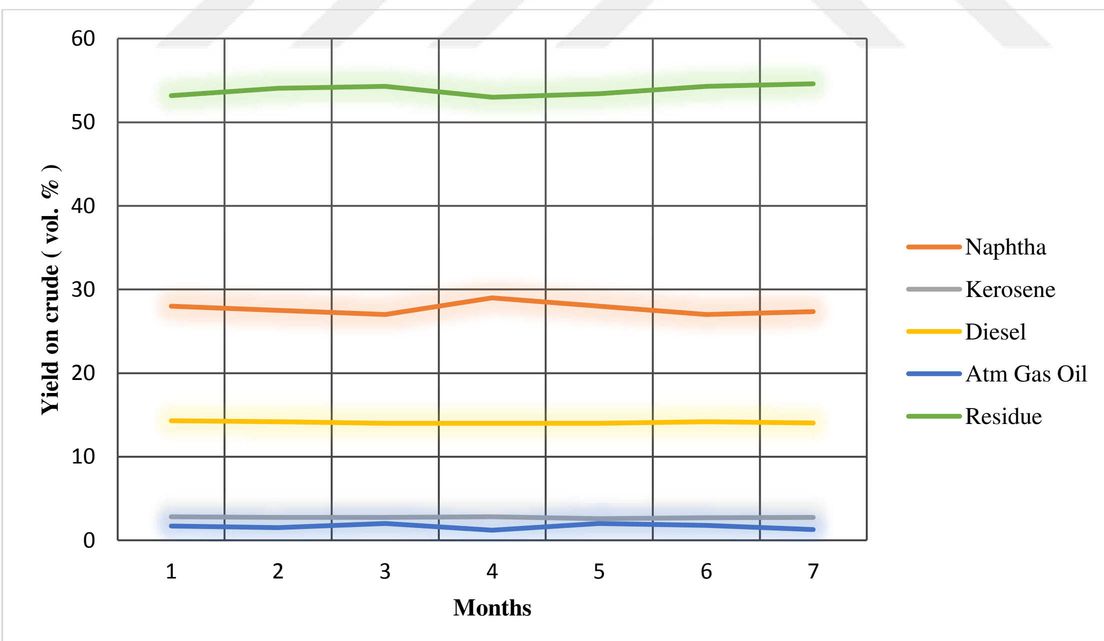


Figure 5.1 Yield percentage summary of the products

5.2 Simulation of Refinery Process Diagram (PFD)

The simulated process flow diagram in Aspen HYSYS is shown in Figure 5.2. The diagram consists of preheat exchanger train. Here, various process streams from crude tower exchange heat with the incoming crude. Typically, the charge stream is heated from 90 °F to 653 °F (32-325 °C).

After leaving the initial preheat exchangers (E-101A/B, -E-109, E-102 and E-110), the crude flows into desalters (PK-102). Inside the desalters, water and impurities that could cause corrosion in process piping and equipment are removed from crude oil. The crude oil leaves the desalters and passes through the reduced crude exchangers (E-103 &E-104), then flows to the charge heaters (-H-101). After leaving the crude charge heaters, the streams enter the crude tower (T-101) at 653 °F (325 °C).

As shown in Figure 5.3 the column consists of 39 stages with a partial condenser, three side strippers and one pump around. The heated crude is sent into the tray 36. Side strippers comprising 3 stages have been utilized for kerosene, diesel and atmospheric gas oil (AGO).

5.2.1 HYSYS Program Validation

The program was validated by comparing CDU products data with the data obtained from HYSYS getting started module. The comparison results showed that there is no significant difference in the liquid volume flow of the products. This comparison indicated that in the present research, the HYSYS program can be safely used to make the retrofit as well as the adaptation measurement as presented in Table 5.2.

Table 5.2 Comparison between started module of HYSYS results and results of this research

Product	Volume flow		Error percentage
	HYSYS results	Research results	
Naphtha	35	34.99998	1.40E-05
Kerosene	7	6.99997	5.32E-05
Diesel	20	20.0001	4.77E-05
Residue	53	52.99999	2.00E-05

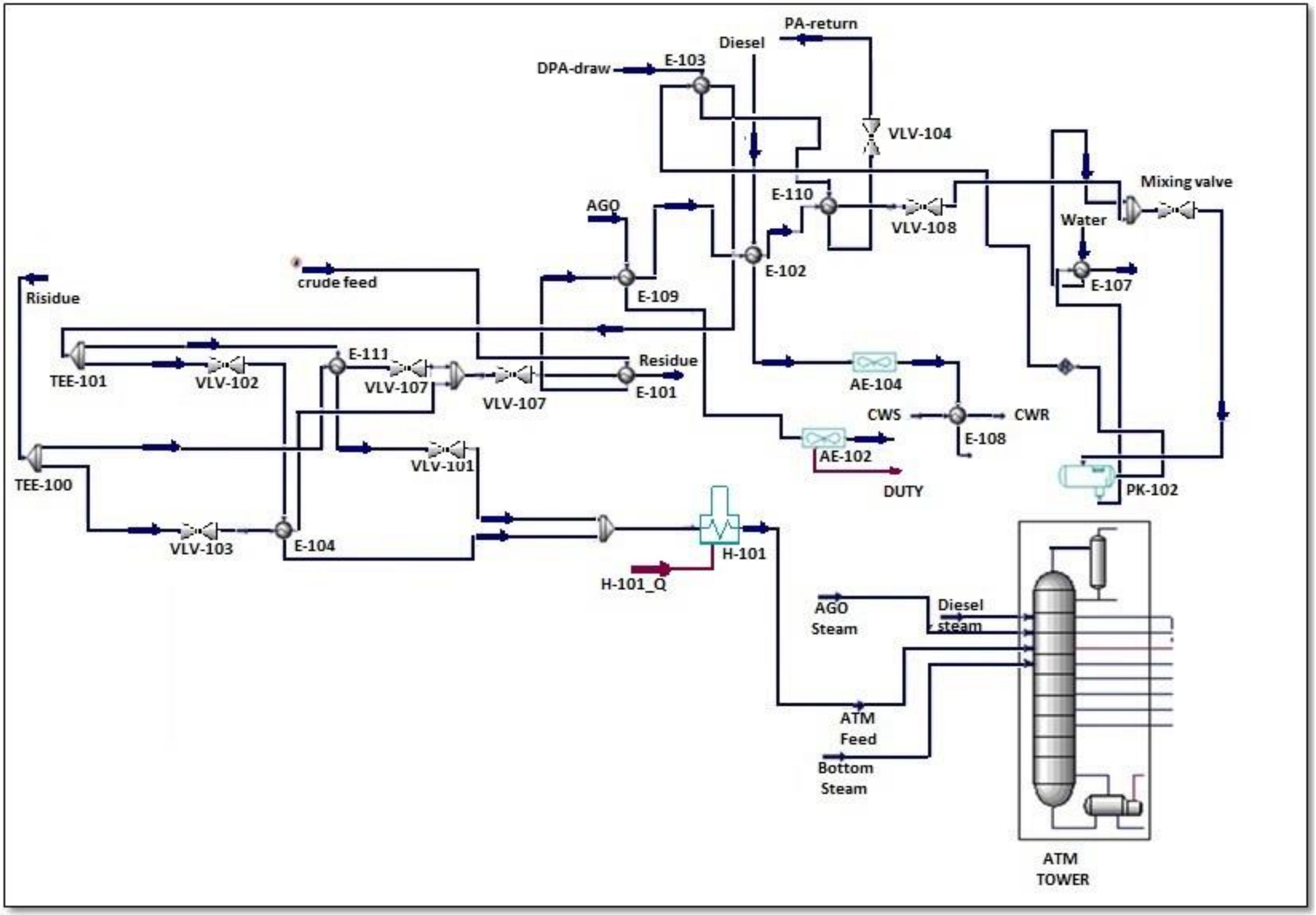


Figure 5.2 Process flow diagram of simulation in HYSYS

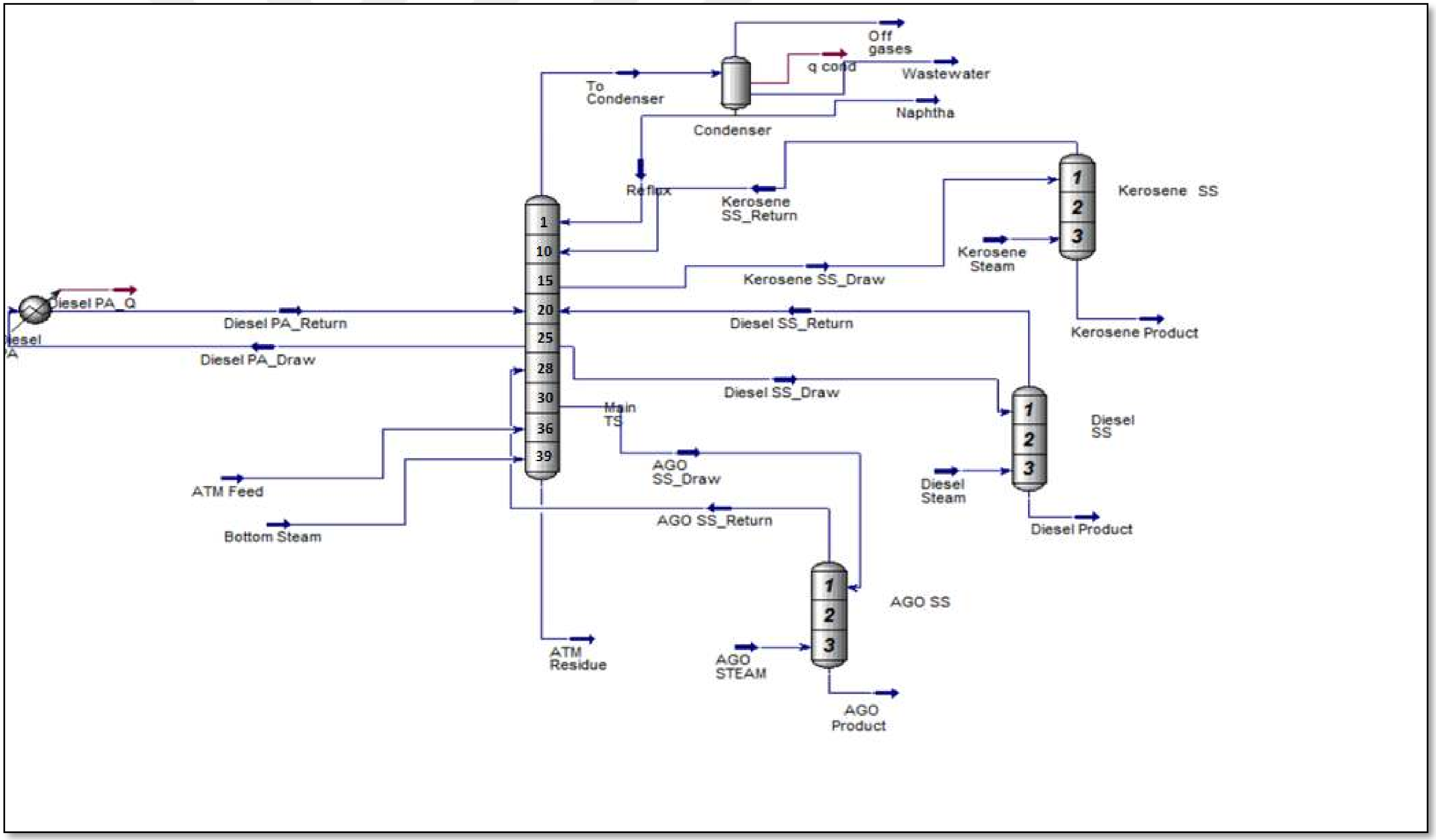


Figure 5.3 Distillation column sub flow sheet in HYSYS

5.3 Effect of Changing Parameters

5.3.1 Influence of increasing Cut Temperature of Kerosene Production

The effect of temperature on the optimization of the products from HYSYS Process modeling software that are of greater commercial value are is presented in Table 5.3.

Table 5.3 The effect of increasing kerosene cut temperature on some variables

Kerosene cut temperature ˚C	Kerosene volume Flow m^3/h	Kerosene Sulfur wt%	Final boiling point kerosene	Diesel sp.gr	Diesel Flow m^3/h
202	3.2	0.21	207	0.824	16.4
202.2	3.3	0.211	207.1	0.82395	16.35
202.4	3.4	0.212	207.2	0.8239	16.3
202.6	3.5	0.213	207.3	0.82385	16.25
202.8	3.6	0.214	207.4	0.8238	16.2
203	3.7	0.215	207.5	0.82375	16.15
203.2	3.8	0.216	207.6	0.8237	16.1
203.4	3.9	0.217	207.7	0.82365	16.05
203.6	4	0.218	207.8	0.8236	16
203.8	4.1	0.219	207.9	0.82355	15.95
204	4.2	0.22	208	0.8235	15.9
204.2	4.3	0.221	208.1	0.82345	15.85
204.4	4.4	0.222	208.2	0.8234	15.8
204.6	4.5	0.223	208.3	0.82335	15.75
204.8	4.6	0.224	208.4	0.8233	15.7
205	4.7	0.225	208.5	0.82325	15.65
205.2	4.8	0.226	208.6	0.8232	15.6
205.4	4.9	0.227	208.7	0.82315	15.55
205.6	5	0.228	208.8	0.8231	15.5
205.8	5.1	0.229	208.9	0.82305	15.45
206	5.2	0.23	209	0.823	15.4

Results showed that the temperature is proportional to a higher performance but sacrificing the performance of the heaviest product. The optimum point was defined by the ASTM D86 curve at different temperatures in each plate. It should be noted that the variables such as tower pressure, feed temperature, water vapor flow to the tower and load were kept constant during the test.

Table 5.3 shows the results obtained for the different cases in which the behavior of some variables and qualities are orientated to produce the maximum product namely kerosene cut.

When the crude oil distillation is done, the temperature is measured in the exit of the kerosene flow. The steams are rising across the trays where the contact is taken place between the vapor and liquid at about 202 °C that is the boiling point of the kerosene at atmospheric pressure. In this respect, the flow that one withdraws from the tower of distillation at this temperature possesses a chemical composition corresponding to the kerosene. The values of all the variables and qualities change as seen in Table 5.3 when the draw temperatures vary in ascending manner.

Making an analysis of the behavior trends, it is possible to observe that as the draw temperature is increased, the flow of kerosene increases too, which is important since the performance of the product increases.

It is obviously seen in Figure 5.4 that the performance of the flow of diesel diminishes with the increase of the cut temperature of the kerosene. This is due to a part of this cut is dragged towards the tray of top retirement due to lager vaporization of heavier components, which increases the flow rate of kerosene product and will affect also the specific gravity of the kerosene due to the content of heavier components in this flow. This situation obeys the balance of mass and energy established.

Having increased the draw temperature one notices that the content of sulfur increases due to the speed in the retirement tray. This causes heavy components to depart from their tray for the tray of the kerosene.

Figure 5.5 shows the analysis of the behavior of the physical/chemical variables. As can be seen in the figure when the plate temperature increases from 202 °C to 206 °C at a rate of 0.2 °C, the flow rate of kerosene increases about 62.5 %, which is important because the performance of the product increases.

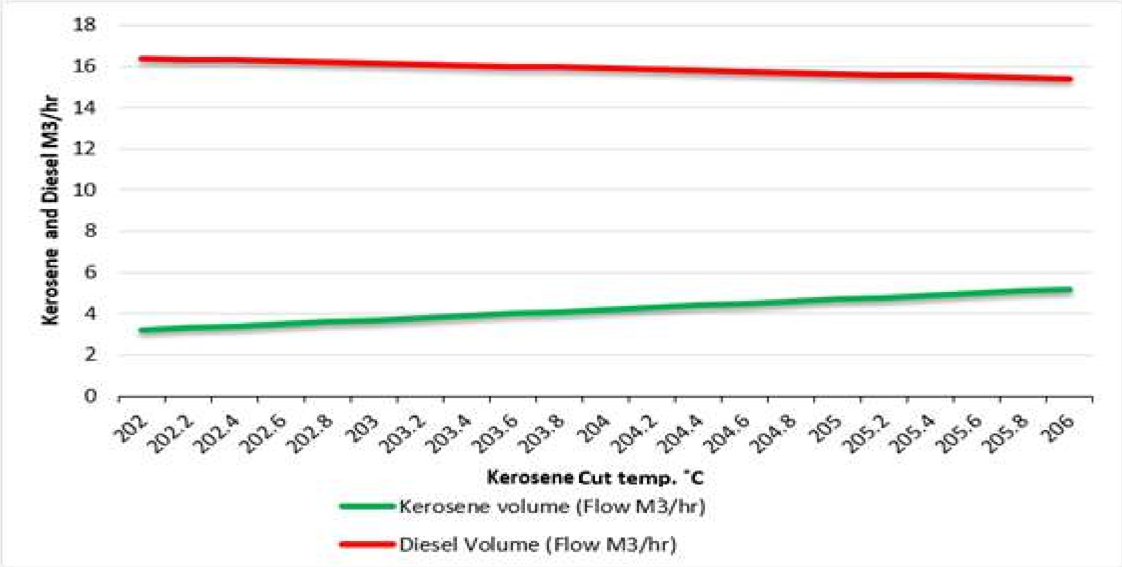


Figure 5.4 Effect of increasing cut temperature of kerosene in kerosene and diesel production flow

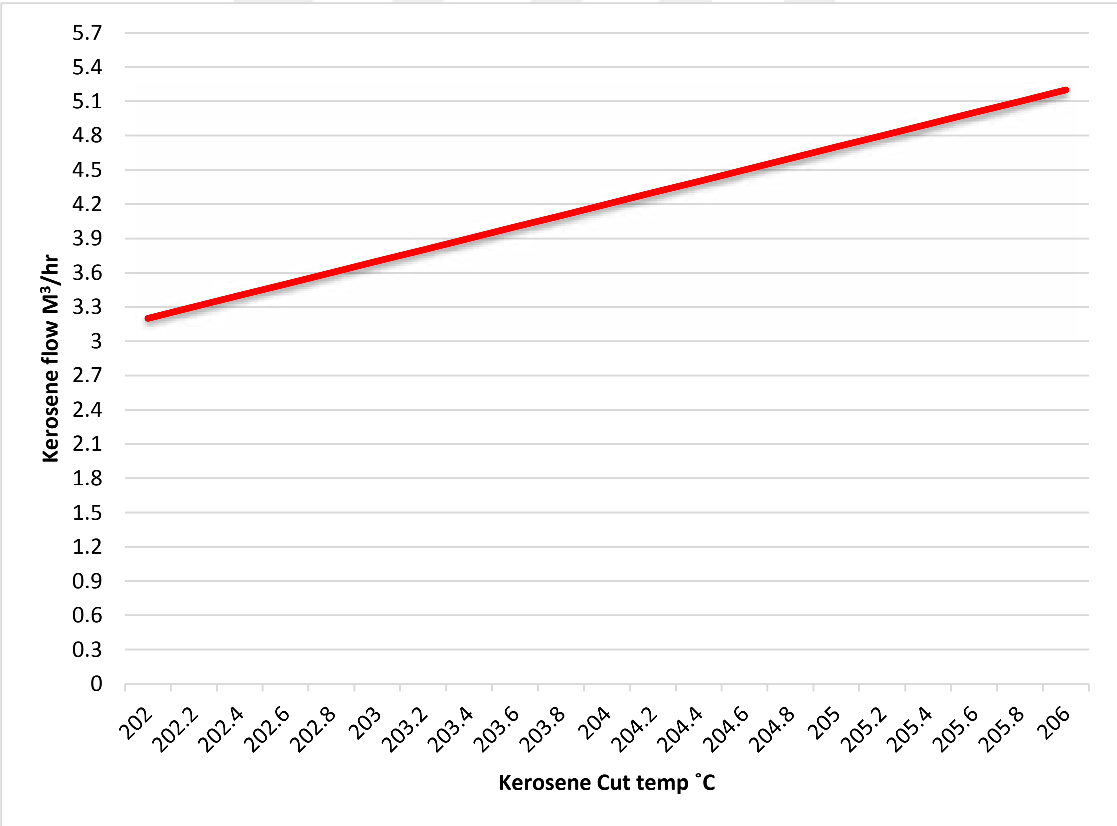


Figure 5.5 Variation of kerosene flow rate with Kerosene Cut temperature

The variation of the sulfur content is directly proportional to increasing temperature of the kerosene plate. The flow rate of heavier components toward to lighter components increases with increasing tray temperature, then sulfur content in kerosene increase since some sulfur will be carried with the heavy components. It is necessary to keep the sulfur content in kerosene under control in a very effective way since it is one of the undesirable components and could cause the loss of quality due to being out of specification in terms of customer requirements. A high concentration of sulfur will produce acidic gases and a high rate of corrosion in the equipment used.

By changing the withdrawal temperature of the dish, the final boiling point of the distillate (kerosene) changes as shown in Figure 5.6, increasing the gap-overlap. When the tendency is to increase the gap, it can be said that the optimization of the product is good because it has a better separation (within 5% of the heavy product and 95% of the light cut), otherwise it will occur if the overlap increases. This is because a part of this cut is dragged towards the upper tray due to the larger evaporation of heavier components, which increases the kerosene production flow and will affect the specific gravity of the kerosene due to the content heavier components in the flow. This situation obeys the established mass and energy balance. An increase in sulfur content in kerosene indirectly affects flow of heavier products such as diesel; therefore, the expenses of treatment to remove sulfur in diesel will be less since flow rate of diesel decreases from 16.4 m^3/h to 15.4 m^3/h. A certain amount of naphtha also changes into kerosene during the handling of the kerosene extraction at plate temperature, which increases flow rate of kerosene.

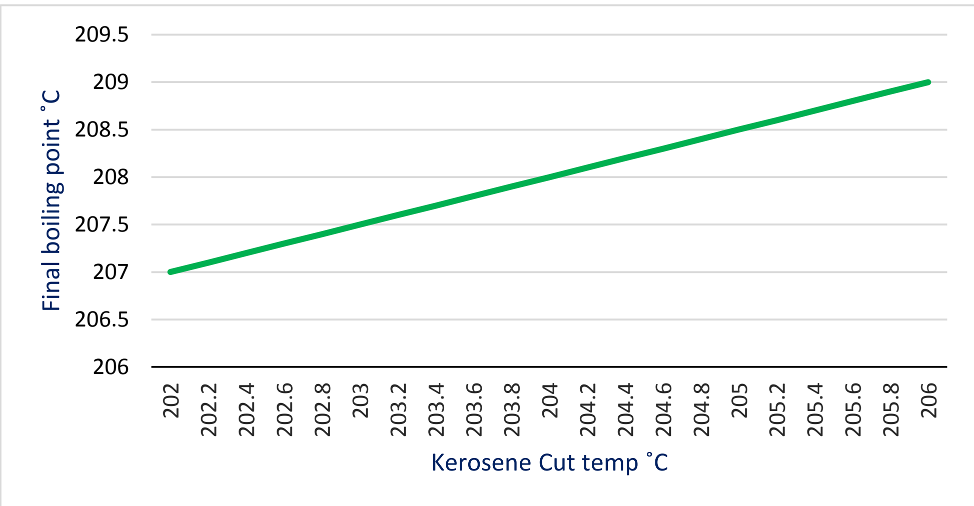


Figure 5.6 Effect of increasing cut temperature of kerosene in kerosene final boiling point

5.3.2 Influence of Decreasing Top Tower Temperature

Decreasing distillation top tower temperature, heavier naphtha products go toward the kerosene, which increases production of kerosene and decreases its distillation end point; therefore, kerosene product becomes lighter. It is visible from Table 5.4.

Table 5.4 Shows the effect of decreasing top tower temperature

Top Temperature ˚C	Kerosene volumetric flow rate m^3/h	Sulfur content wt %	Final boiling point kerosene	Naphtha flow m^3/h	Naphtha API	Naphtha RVP Psi
153	3.2	0.21	207	32	64.8	7.2
152.8	3.3	0.209	206.8	31.9	64.82	7.3
152.6	3.4	0.208	206.6	31.8	64.84	7.4
152.4	3.5	0.207	206.4	31.7	64.86	7.5
152.2	3.6	0.206	206.2	31.6	64.88	7.6
152	3.7	0.205	206	31.5	64.9	7.7
151.8	3.8	0.204	205.8	31.4	64.92	7.8
151.6	3.9	0.203	205.6	31.3	64.94	7.9
151.4	4	0.202	205.4	31.2	64.96	8
151.2	4.1	0.201	205.2	31.1	64.98	8.1
151	4.2	0.2	205	31	65	8.2
150.8	4.3	0.199	204.8	30.9	65.02	8.3
150.6	4.4	0.198	204.6	30.8	65.04	8.4
150.4	4.5	0.197	204.4	30.7	65.06	8.5
150.2	4.6	0.196	204.2	30.6	65.08	8.6
150	4.7	0.195	204	30.5	65.1	8.7

As can be seen in Table 5.4, the kerosene flow rate increases as the uppercut temperature decreases (product with lighter naphtha components). This naphtha becomes a part of the kerosene flow; therefore, the kerosene yield increases. it can be noted that the sulfur content in kerosene cutting decreases with a decrease in the upper cutting temperature. This is because the higher product returning to the kerosene has a very low sulfur content;

the sulfur tends to remain in the heavier cuts such as light diesel and diesel. The lowering of the temperature in the upper plate of the distillation tower causes a variation in the end point of the kerosene as part of the light product remains in the kerosene cut, which causes the final boiling point of the kerosene to decrease since it has a higher content of low molecular weight carbonate chains. On the other hand, it is observed that the flow rate of naphtha is directly proportional to the temperature variations of the upper cut. In present case, its decrease will produce a smaller extraction of this cut, which decreases its yield and increases the yield of the kerosene production. This situation complies with the mass and energy balance. Thus, having a lower molecular weight liquid current returning to the kerosene plate will generate a higher API.

When the temperature of the naphtha cut decreases, its heavier fractions remain in the lower tray so that the lighter product contents increase their concentration, producing an increase in the RVP in the naphtha.

5.3.3 Influence of Increasing Steam Flow Rate

Increasing mass flow rate of steam to distillation tower, partial pressures of products decrease and heavier products will go up and also the percentages of sulfur in the cuts will increase because of dragging of it with heavier products.

As it is obvious from Table 5.6, the stripping steam under conditions of 300˚C and 14 kg/m^2 in a range of 700 kg/h - 775 kg/h is supplied at the bottom of the distillation tower. The final boiling point of the cut increases as the steam flow at the bottom of the distillation tower increases. The heavy components will be dragged to the upper tray since the surface tension of the component mixture decreases. Heavier components contain more sulfur because they are more similar, so the sulfur content also increases.

An increase in upstream steam traffic because of increased steam flow rate is equal to an increase in kerosene cutting production. In addition, the rising vapors carry light components to the upper trays, which will condense according to their partial pressure at the tray temperature. Thus, the performance of the naphtha will increase. The increase in the production of the different cuts such as naphtha and kerosene is due to the light components overlapped in the residue, but they also carry by dragging heavy components having a high sulfur content.

Table 5.5 The effect of increasing steam flow rate on some variables

Steam kg/h	Final boiling point kerosene	Kerosene Sulfur content wt%	Kerosene volume flow rate m^3/h	Final boiling point naphtha	Naphtha volume flow rate m^3/h	Naphtha RVP Psi
700	207	0.21	3.2	189	32	7.2
705	207.15	0.2104	3.3	189.35	32.2	7.3
710	207.3	0.2108	3.4	189.7	32.4	7.4
715	207.45	0.2112	3.5	190.05	32.6	7.5
720	207.6	0.2116	3.6	190.4	32.8	7.6
725	207.75	0.212	3.7	190.75	33	7.7
730	207.9	0.2124	3.8	191.1	33.2	7.8
735	208.05	0.2128	3.9	191.45	33.4	7.9
740	208.2	0.2132	4	191.8	33.6	8
745	208.35	0.2136	4.1	192.15	33.8	8.1
750	208.5	0.214	4.2	192.5	34	8.2
755	208.65	0.2144	4.3	192.85	34.2	8.3
760	208.8	0.2148	4.4	193.2	34.4	8.4
765	208.95	0.2152	4.5	193.55	34.6	8.5
770	209.1	0.2156	4.6	193.9	34.8	8.6
775	209.25	0.216	4.7	194.25	35	8.7

5.3.4 Influence of Increasing Crude Feed Temperature

Increasing temperature of the crude oil fed to distillation tower increases the rate of process separation of the product and decreases time of process but increases the pressure of the tower. This causes high boiling point components to start boiling and thus, heavier products that will go up in the tower affect the quality of the products. It is noticeable in Table 5.6.

Table 5.6 The effect of increasing crude feed temperature some variables

Feed temperature ˚C	Kerosene volumetric Flow rates m^3/h	Sulfur content wt %	Final Boiling Point of kerosene	Naphtha sp.gr	Naphtha volumetric Flow rates m^3/h	Naphtha RVP Psi	Final Boiling Point of naphtha
325	3.2	0.21	207	0.7203	32	7.2	189
325.2	3.3	0.211	207.23	0.7204	32.3	7.28	189.2
326	3.4	0.212	207.46	0.7205	32.6	7.36	189.4
326.5	3.5	0.213	207.69	0.7206	32.9	7.44	189.6
327	3.6	0.214	207.92	0.7207	33.2	7.52	189.8
327.5	3.7	0.215	208.15	0.7208	33.5	7.6	190
328	3.8	0.216	208.38	0.7209	33.8	7.68	190.2
328.5	3.9	0.217	208.61	0.721	34.1	7.76	190.4
329	4	0.218	208.84	0.7211	34.4	7.84	190.6
329.5	4.1	0.219	209.07	0.7212	34.7	7.92	190.8
330	4.2	0.22	209.3	0.7213	35	8	191
330.5	4.3	0.221	209.53	0.7214	35.3	8.08	191.2
331	4.4	0.222	209.76	0.7215	35.6	8.16	191.4
331.5	4.5	0.223	209.99	0.7216	35.9	8.24	191.6
332	4.6	0.224	210.22	0.7217	36.2	8.32	191.8
332.5	4.7	0.225	210.45	0.7218	36.5	8.4	192

The high temperature of feed increases the flow rate of higher molecular weight elements from the flash or feed plate to the grinding area, mixing with the lighter fraction and increasing its final boiling point.

Higher specific gravity corresponds to higher molecular weight elements, which indicates that the increase in the specific gravity due to higher feed temperature is results of the heavier compounds that flow to higher elevations and thus change the composition of light products. This condition generates a higher yield of products such as naphtha and kerosene, which results in an increase in sulfur content and higher specific gravity and final boiling point.

Higher temperature in the flash tray (feeding the tray) sends heavier components to the upper tray; therefore, a part of the sulfur content that should be deposited in the diesel dish is transferred to the upper cut, which causes an increase in sulfur content in the kerosene. This can be controlled at the request of the specifications demanded by the client.

5.3.5 Influence of Decreasing Tower Pressure

A decrease in the partial pressures of products with decreasing pressure at the top of the tower increases flow rates of cut production; however, the quality of the products decreases since sulfur percentage increases as evidenced in Table 5.7.

Table 5.7 The effect of decreasing tower pressure on some variables

Column Pressure Bar	Kerosene volume Flow m^3/h	Sulfur content wt %	Final Boiling Point kerosene	Naphtha sp.gr	Naphtha volume Flow m^3/h	Naphtha RVP Psi
1.7	3.2	0.21	207	0.7203	32	7.2
1.685	3.4	0.2111	207.24	0.7205	32.4	7.18
1.67	3.6	0.2122	207.48	0.7207	32.8	7.16
1.655	3.8	0.2133	207.72	0.7209	33.2	7.14
1.64	4	0.2144	207.96	0.7211	33.6	7.12
1.625	4.2	0.2155	208.2	0.7213	34	7.1
1.61	4.4	0.2166	208.44	0.7215	34.4	7.08
1.595	4.6	0.2177	208.68	0.7217	34.8	7.06
1.58	4.8	0.2188	208.92	0.7219	35.2	7.04
1.565	5	0.2199	209.16	0.7221	35.6	7.02
1.55	5.2	0.221	209.4	0.7223	36	7
1.535	5.4	0.2221	209.64	0.7225	36.4	6.98
1.52	5.6	0.2232	209.88	0.7227	36.8	6.96
1.505	5.8	0.2243	210.12	0.7229	37.2	6.94
1.49	6	0.2254	210.36	0.7231	37.6	6.92
1.475	6.2	0.2265	210.6	0.7233	38	6.9

As the pressure in the tower decreases, the separation of the products tends to improve and thus, the volumes of gases increases. Therefore, lighter cuts such as kerosene and naphtha tend to increase their production. The difficulty of separation decreases with increasing relative volatility. Thereby, the number of floors, reflux and consumption requirements in the condenser and the boiler decrease.

The effect of the lowering of the tower pressure on the sulfur content in kerosene cutting was examined by reducing tower pressure while other parameters were kept to be constant. As seen in Table 5.8, the sulfur content in kerosene increases with the reduction in tower pressure as the gas flow traffic increases and a part of the lower product (diesel) with its sulfur load flows into the upper product namely kerosene. As the upper pressure of the tower decreases, upper products increase due to the mobilization of heavy carbon chains to the upper plates, it is noted that a heavier cut will have a higher final boiling point.

The specific gravity of a raw cut depends on its components. The specific gravity of kerosene increases due to a part of diesel in the kerosene since higher molecular weight causes the higher specific gravity in a mixture.

In the comparative Table 5.8, it can be noted that the least favorable action for the process is to increase the temperature of feed. This is because besides the increased sulfur concentration; the specific gravity and the final boiling point can increase.

The specifications requested by the customer generates a higher consumption of fuel oil in the furnaces to reach this temperature, which requires a higher operating cost in the use of fuel oil whose volumetric flow rate increases by $0.23m^3/h$ for each additional burner that is require. On the other hand, the most recommended action is to lower the top pressure since it does not require an increase in operating expenses and maximizes the production of light cuts (kerosene and naphtha). However, attention must be paid to the concentration of sulfur in kerosene that must not exceed the required specifications for the consumption/customers.

Results indicate that there is high variation in volume percentage yield between Erbil (refinery) and simulation by HYSYS, which indicate high losses in light product yield. Table 5.9 shows that the volume flow rate of light products are greater in the simulation than in Erbil refinery where is the heavy product (residue) in HYSYS is less than in the refinery.

Table 5.8 Behavior of the optimized variables in the T-101 distillation tower

Manipulated variable	**Variation**	**Kerosene volume**	**sulfur**	**final boiling point**	**Specific gravity**	**Naphtha volume**
Cut temperature	↑	↑	↑	↑	↑	↓
Top tower temperature	↓	↑	↓	↓	↓	↓
Steam flow rate	↑	↑	↑	↑	↑	↑
Feed temperature	↑	↑	↑	↑	↑	↑
Column pressure	↓	↑	↑	↑	↑	↑

5.4 Comparison Between HYSYS Simulation and Erbil Refinery Results

The table allows to see the differences between the case study and the real data from the refinery, results of HYSYS vs. results being applied in the Erbil refinery. In principle, it is noted that all the yields of the cuts having high value in terms of price increase while yields of cuts such residue having less value decrease. It is can be also noticed that the kerosene cut is the one that presents a greater increase in efficiency. It is due to the fact that the actions taken are designed with the purpose of increasing flow rate of kerosene. Rising the temperature of the kerosene that is important parameter will obviously have a greater recovery of this current. With the drag of more flow to the top of the distillation column, all cuts will have an addition because the components lighter than the fuel will flow to the top. The difference between the theoretical behavior (HYSYS) and the practical (Experimental) is due to the variations in the process parameters such as temperature and tower pressure.

In the experimental case, the crude oil characteristics can vary since it comes from different productive fields. Besides being in the storage tank, this is stratified and may be sending to plant a lighter or heavier crude at some points.

Table 5.9 Comparison of values between HYSYS simulation and real results in the refinery

Product	unit	Erbil refinery	HYSYS
Naphtha flow rate	m^3/h	32	35
Kerosene flow rate	m^3/h	3.2	7
Diesel flow rate	m^3/h	16.4	20
AGO flow rate	m^3/h	1.5	2
Fuel flow rate	m^3/h	63.87	53

The equipments that make up the distillation unit have fouling and heat exchanger losses, which lowers the efficiencies to other possible internal leaks in the pipes. The deviations in the measurements of the instruments of pressure, level and temperature are being controlled and it is seen that the control of those variables is not entirely stable and a slight deviation is allowed all this adds up. Whereas in the simulation, the characteristics of the oil are invariable and its behavior responds to equations. The measurements indicated that the simulated instruments are the result, without deviation, of the values fed to the system; there are no instrument failures or presence of water in low points that could be several values for the pressure and temperature of the currents in studio.

The decrease of the waste flow to maximize the flow of the cuts of interest is an achievable ideal but in any case, it will depend on the characteristics of the crude oil and the operation of the distillation unit and is affected by the intricate variables of the process.

5.4.1 Properties of Kerosene before and after Optimizing

The results of kerosene, which is provided by HYSYS program is compared to the base case of the refinery as shown in Table 5.10 that shows an increase in the values of some properties of kerosene. This is because a greater flow rate of output for kerosene due to the increase of the feed temperature is left in this cut because of more heavy components.

As can be seen in Table 5.10, when the specific gravity increases from 0.7898 to 0.7943, the flow rate of ideal liquid change from 3.2 m^3/h to 7 m^3/ h.

Table 5.10 Properties of kerosene

Properties	Before Optimizing	After Optimizing
Std Ideal Liq Vol Flow (m^3/h)	3.2	7
SP.Gr	0.7898	0.7943
Total Sulfur wt %	0.21	0.2234
Molecular Weight	132.4	135.4
Viscosity (CP)	0.201	0.2146

Sulfur is a compound that tends to stay in the heavier cuts; therefore, since more diesel goes to kerosene, the more sulfur will carry to kerosene with it. Thereby, this explains the reason why the sulfur content increases in kerosene as evidenced in Table 5.10. Since the molecular weight of a compound is known as mass in a mole of that compound, the larger carbon chains present in the stream results in larger the molecular weight. Understanding the simulation, it will be noticed that the heavier hydrocarbon compounds are flowing the kerosene, which some chains belonged to diesel (overlap) are carried to the kerosene and make kerosene cut heavier. Technically speaking, viscosity can be defined as the opposition of a fluid flow to tangential deformations. While the molecules are larger, their cohesion forces will be greater since the molecular orientation and cohesion forces of the fluid increase with increasing molecular weight and thus, their viscosity increases. That is why kerosene with a higher molecular weight leads to a higher viscosity, as shown in the results.

5.4.2 Optimum Operation Conditions Obtained by HYSYS

The summary of optimum results comparing the base and optimized cases of the model is presented in Table 5.11. The optimized case showed improvements in naphtha and kerosene productions with volumetric flow rates of 35 m^3/h and 7 m^3/h over the base case, respectively.

By observing and taking into consideration the results shown in Table 5.11, the most important change made to obtain the optimization was the increase in the transfer temperature.

Table 5.11 Summary of optimized cases and optimum results of the base case

Parameters	Units	Base case	Optimized case
Crude Feed Flow rate	m^3/h	117	117
Heater temperature	˚C	325	325
Column top temperature	˚C	153	157
Column top pressure	bar	1.7	1.5
Steam Flow rate	Kg/h	700	750
Kerosene Cut of temp.	˚C	202	206
Diesel Cut of temp.	˚C	251	260

It will be understood that this does not generate any change in the flow of feed; therefore, this remains the same. The variations in quantities of the steam flow, the kerosene and diesel cutting temperatures help to break the van der Waals forces. With this, it improves the stripping of the lighter compounds of the crude oil, which help to increase the flow rate of lighter compounds from lower tray to upper tray of the distiller; therefore, in order to increase the recovery of kerosene, the steam stripping has to be increased.

The temperature of the kerosene tray presents a slight increase in flow rate of kerosene, which indicates that this stream has to be withdrawn as much as possible. An increase few degrees in temperature help the yield to increase. The fact of having a slightly heavier composition means the increases of end boiling point; therefore, few degrees raised in the temperature is very helpful to remove stream in the tray.

An increase in the temperature of the diesel tray cause an increase in flow of the lighter components from diesel tray to kerosene tray. Thus, the end point of this stream rises, which means better stripping and results in greater flow of ascending vapors that will be condensed in the upper trays, are mostly in the kerosene and other in naphtha stream.

6. CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusion

In this study, crude distillation unit was simulated and verified using HYSYS simulation program to analyze the influence of variation of some parameters on products. In other words, the optimum operation conditions were determined to obtain maximum kerosene. This program was used to study to determine the effects of varied temperature, pressure, and flow rates on the products. The results indicated that the product yields were not stable and often changed according to the variation of operation parameters.

The simulation results were compared with Erbil refinery results. The analysis showed there have been several variations in yields of products between HYSYS simulation and Erbil refinery results. However, the flow rate of naphtha in HYSYS was continuously greater than that in Erbil refinery in which flow rate of naphtha may vary according to crude oil composition, whereas for residue the yield in HYSYS was less than that in Erbil refinery.

In today's energy use, crisis of low production rate against the increasing demand of the fuel production regularly hampers both the domestic and industrial operations since naphtha and kerosene are the major power source of this country. Unless other power source is developed, the refinery productions are our only hope. Almost all the existing processing plants are now operating beyond their capacities. Nonetheless, there has been a dwindling situation in the naphtha and kerosene production. In such a case, an idea of optimization of the distillation processing plant will surely pave a way to a sustainable solution. The steady state simulation of the naphtha and kerosene processing plant was performed based on the design and physical properties of those compounds.

Aspen HYSYS is a process simulation environment designed to serve many processing industries. It is an interactive, intuitive, open and extensible program. It also has many add on options to extend its capabilities into specific industries. Rigorous steady state and dynamic models for a plant design can be created with this program.

Apart from this, monitoring, troubleshooting, operational improvement, business planning and asset management can be performed with the HYSYS process simulator.

Through the completely interactive HYSYS interface, process variables and unit operation topology can be easily manipulated (Aspen, 2017).

All distillation columns ought to be rigorously operated to attain the specified production rates and products quality. The three main objectives of column management are declared as:

- To set stable conditions for column operation.
- To regulate conditions within the column in order that the product(s) continuously to meet the specified.
- Specifications to attain the on top of objective most with efficiency, by increasing product yield, minimizing energy consumption, etc.

Process variables like temperatures, pressures, flow rates, levels and compositions should be monitored and controlled altogether in distillation processes. These process variables in a distillation system have an effect on one another, whereby a modification in one process variable can lead to changes in different process variables. Thus, in column management one ought to be watching the entire column and not that specialized in any specific sections solely.

Each column contains a system that consists of many management loops. The loops regulate process variables required to catch up on changes because of disturbances throughout plant operation. Each process variable has its own management loop, which usually consists of a detector and transmitter, controller and control valve. Each control loop keeps track of the associated process variable. An adjustment is made to a process variable by varying the opening of its control valve. The stream flow rate is, therefore, adjusted and a desirable variable is being controlled.

- Column pressure commonly controlled at a constant value,
- Feed rate typically set by the amount controller on a preceding column,
- Feed rate is severally controlled if fed from tank,
- Feed temperature controlled by a feed preheater. Before preheater, feed could also be heated by bottom product via feed/bottom exchanger,
- Top temperature sometimes controlled by varied the reflux,
- Bottom temperature controlled by varied the steam to reboiler,
- The compositions controlled by regulation the reflux flow and boiled-up,

- Pressure is commonly thought of the prime distillation management variable because it affects temperature, condensation, vaporization, compositions and volatilities for nearly any process that takes place within the column. Column pressure management is often integrated with the condenser system. Reboilers and condensers are an integral part of a distillation system. They regulate the energy in and outflow in a very distillation column. A column is controlled by regulation of its material and energy balances.

To achieve the optimization of the kerosene flow rate in the distillation column of the refinery, Erbil, the HYSYS process simulator was used to estimate some variables such as draw temperature, top tower temperature, feed temperature, top tower pressure and steam flow.

The results of simulation indicated that residue product had the highest volumetric flow rate, while kerosene and AGO had the lowest value of volumetric flow rate. This showed that the optimization of the column is required to convert more of the atmospheric residue into other premium products like naphtha, kerosene and diesel.

In summary, after reviewing the current operation conditions in Erbil refinery and considering the main operation parameters used in the Aspen HYSYS process simulator, it was possible to find the optimum operating conditions that maximize the production of naphtha and kerosene for a certain crude oil. The simulation indicates that the flow rates of the naphtha and kerosene can be increased to be from 27.36% to 29.91% and from 2.74% to 5.98%, respectively. Both simulation and optimization were tested for different operating conditions and it is observed that program achieves a rapid convergence of the used models. This leads to the same models can be applied to other distillation towers with different designs.

6.2 Recommendations

Following recommendations are put forward for future work:

- Use Aspen HYSYS to all other units in the refinery for optimization.
- Make simulation and optimization for modifying original design conditions. Perform cost analysis for future work.
- Use another simulation software such as Aspen Plus, CHEMCAD, ProModel, etc. in order to compare its results with the current results and find out differences among data obtained different simulation program for the same variable.

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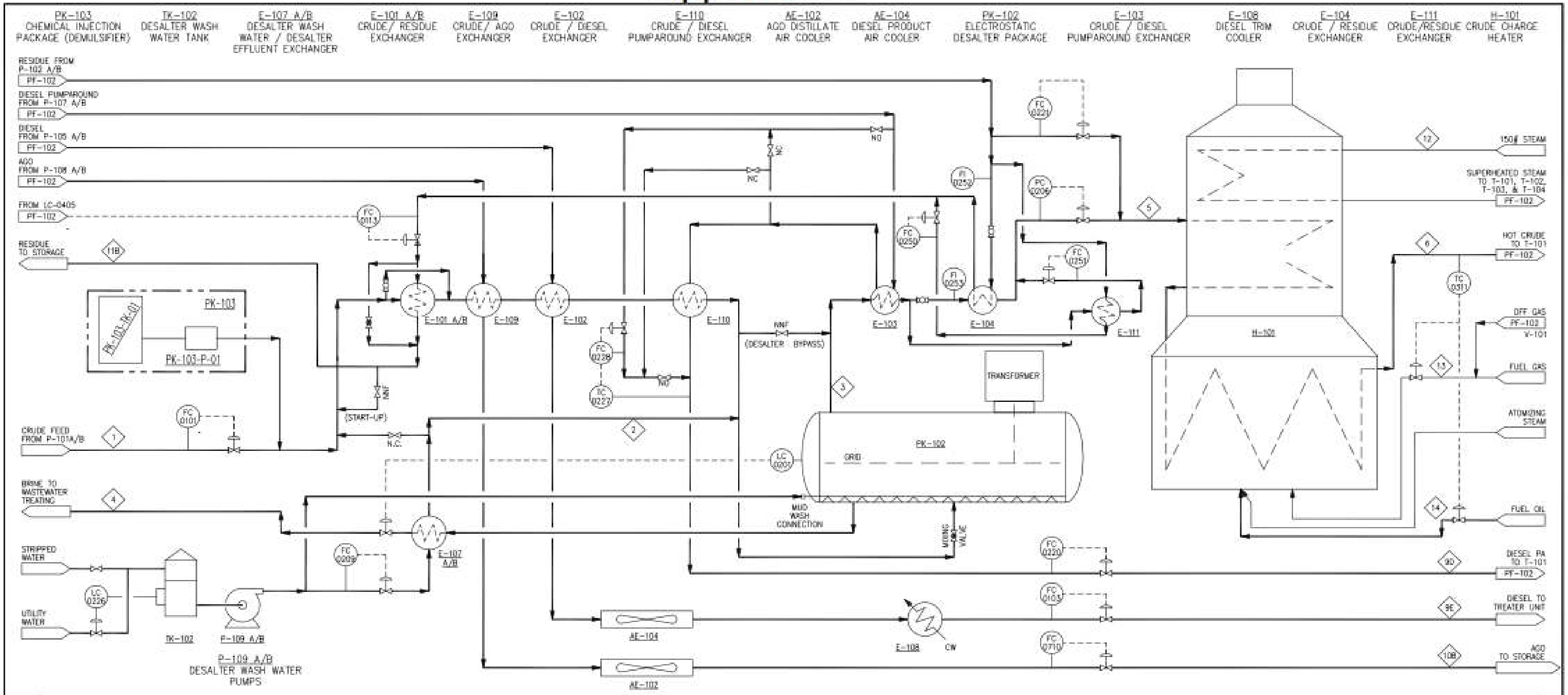
APPENDICES

APPENDIX A (PFD)

APPENDIX B (Operational Conditions and Laboratory Results, Design Specification)

APPENDIX C (HYSYS Reports)

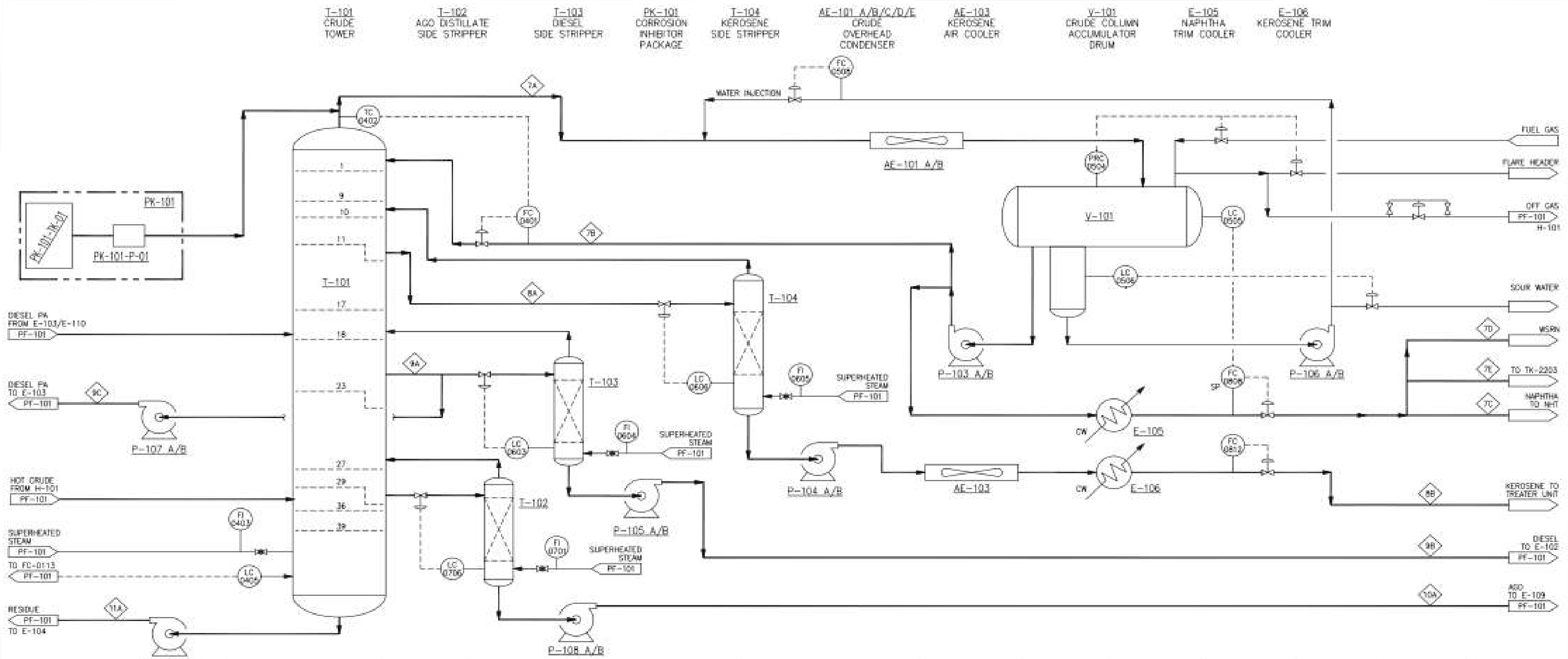
Appendix A



Stream Number	1	2	3	4	5	6	9D	9E	10B	11B	12	13	14
Description	Raw Crude from Tankage	Washwater to Desalter	Desalted Crude from PK-102	Water Effluent from E-107	H-101 Inlet	H-101 Outlet	Diesel PA	Diesel to Treater	AGO Product Rundown	Residue to Storage	150# Steam to H-101	Fuel Gas to H-101	Fuel Oil to H-101
Phase	L	L	L	L	M (V/L)	M (V/L)	L	L	L	L	V	V	L
Temperature, °C	37.8	109.4	138	60.0	196	354.4	187.8	43.3	54.4	90	197.8	37.8	85.0
Pressure, kg/cm2	16.5	12.7	11.0	4.9	8.6	2.5	2.1	7.0	6.5	3.4	10.5	7.0	10.5
Liquid M3/H Actual Condition	135	6.6	132.7	8.0			40.5	31.0	4.0	49.0			
Vapor Normal M3/HR Actual Condition													
Mass Flow, kg/hr	111,615	7,923	111,842	17,000	111,842	111,842	28,167.5	25,000	4,140	46,053	1,533.4	1,700	2,100
Molecular Weight	175.5	18.0	172.2	18.0	64/183	125/338	192	272	282	370	18.0		
Enthalpy,Kjoule/kg	-2126.3	-2077.0	-1979.5	-1973.7	-1806.3	-1185.9	-1780.5	-2099.7	-1386.3	-2057.7	-6478.0		
Viscosity, cp	5.29	0.6	0.74	0.3	0.01/0.39	0.01/0.26	0.3366	2.0	9.2	29	0.1		

CRUDE UNIT
PROCESS FLOW DIAGRAM
DRAWING NUMBER
PF 101

Appendix A (continued)



Stream Number	7A	7B	7C	7D	7E	8A	8B	9A	9B	9C	10A	11A
Description	T-101 Overhead Vapor	Naphtha Reflux	WSRN to NHT	WSRN to E-304	WSRN to Tankage	Kero Draw Off	Kero to Treater	Diesel Draw Off	Diesel to E-102	Diesel PA	AGO to E-109	Residue to E-104
Phase	V	L	L	L	L	L	L	L	L	L	L	L
Temperature, °C	156	70	46.1	46.1	46.1	200	43.3	270	248	270	321	340
Pressure, kg/cm2	2.0	7.0	6.3	6.3	6.3	2.0	9.1	2.1	11.5	2.1	8.2	8.4
Liquid M3/H Actual Condition		89	41	41	41	12.5	10	84.6	38.6	46	6.0	59.0
Vapor Normal M3/HR Actual Condition												
Mass Flow, kg/hr	87,300	124,700	27,250	27,250	27,250	7,900	7,900	53,705	25,050	28,655	4,147	46,136
Molecular Weight	82.0	91	91	91	91	142.5	142.5	202	202	202	282	370.3
Enthalpy,Kjoule/kg	-1799.5	-2125.1	-2200.7	-2200.7	-2200.7	-1681.9	-2142.8	-1531.7	-1629.8	-1531.7	-1384.9	-1337.0
Viscosity, cp	0.0	0.3	0.3	0.3	0.3	0.2	0.94	0.2	0.2245	0.1657	0.1198	0.34

P-107 A/B
DIESEL PUMPAROUND
PUMPS

P-102 A/B
ATMOSPHERIC
RESIDUE
PRODUCT PUMP

P-108 A/B
GAS OIL
PRODUCT PUMPS

P-105 A/B
DIESEL
PRODUCT PUMPS

P-104 A/B
KEROSENE
PRODUCT PUMPS

P-103 A/B
WSRN PRODUCT
AND REFLUX PUMPS

P-106 A/B
SOUR WASH WATER
PUMPS

CRUDE UNIT
PROCESS FLOW DIAGRAM

DRAWING NUMBER
PF-102

Appendix B

Composition Mole Fraction

1	Ethane	0.0047
2	Propane	0.0239
3	i-Butane	0.0145
4	n-Butane	0.0305
5	i-Pentane	0.0326
6	n-Pentane	0.0321
7	H2O	0.0000
8	Air	0.0000
9	H2S	0.0000
10	NBP_44	0.0130
11	NBP_58	0.0296
12	NBP_71	0.0343
13	NBP_86	0.0607
14	NBP_98	0.0373
15	NBP_112	0.0265
16	NBP_129	0.0325
17	NBP_141	0.0476
18	NBP_155	0.0409
19	NBP_169	0.0371
20	NBP_183	0.0348
21	NBP_197	0.0317
22	NBP_211	0.0290
23	NBP_224	0.0268
24	NBP_238	0.0250

25	NBP_252	0.0234
26	NBP_266	0.0221
27	NBP_280	0.0211
28	NBP_294	0.0205
29	NBP_308	0.0202
30	NBP_322	0.0193
31	NBP_336	0.0179
32	NBP_350	0.0164
33	NBP_364	0.0152
34	NBP_378	0.0145
35	NBP_392	0.0141
36	NBP_406	0.0133
37	NBP_420	0.0124
38	NBP_441	0.0222
39	NBP_468	0.0188
40	NBP_496	0.0172
41	NBP_524	0.0128
42	NBP_550	0.0123
43	NBP_579	0.0085
44	NBP_608	0.0083
45	NBP_634	0.0061
46	NBP_677	0.0084
47	NBP_731	0.0048
48	NBP_792	0.0028
49	NBP_850	0.0026
Total		1.0000

ASSAY %	DENSITY Kg/hr
9.23	650.1891
20.76	734.9183
28.82	782.3227
38.11	801.9831
47.55	833.6193
55.68	855.076
62.49	880.8241

ASSAY %	TEMP ˚C
9.23	80
20.76	150
28.82	193
38.11	246
47.55	300
55.68	343
62.49	382

Standard Density kg/m3	843.7686
Viscosity 1 Temp ˚C	40
Viscosity2 Temp ˚C	98.88889

LIGHT END	COMPOSITION
Ethane	0.08
Propane	0.60
i-Butane	0.48
n-Butane	1.01
i-Pentane	1.34
n-Pentane	1.32
H2O	0.00
Air	0.00
H2S	0.00

Appendix B (contiued)

Specification for Simulation

Table B.1 feed operating conditions

Temperature	354 ˚ C
Flow rate	132.5 m3/h

Table B.2 Stripper specification

Stripper	type	Draw Stage	Return Stage
Kerosene	Steam	12	9
diesel	Steam	24	17
AGO	Steam	29	27

Table A.3 Distillation column operating conditions

	Temprature	Pressure kg/cm2
TOP STAGE	155	2
BOTTOM STAGE	340	8.4

Table B.3 Pumps abound operating conditions

	Draw Stage	Return Stage	T1 ˚ C	T2 ˚C
Disel Pump Around	24	18	270	187.8

Table B.4 production data

Naphtha	35
Kerosene	10
Diesel	31
Ago	4
Residue	59

Operation Conditions

Feed Rate M^3/h	H-101 Outlet T ºC	Top T-101 (ºC)	Pre T-101 kg/cm^2	Steam To T-101 Kg/h
177	325	153	1.7	700

Lab Results	Flash ºC (RVP) Psi	Sulfur %wt	Draw Temp. (ºC) Set Point	Flow M3/Hr	Yield vol %
Naphtha	20		153	32	27.36
Kerosene	43	0.21	202	3.2	2.74
Diesel	77	0.7542	251	16.4	14.02
AGO	> 85		295	1.5	1.28
Residue	> 85		310	63.87	54.6

Appendix C

aspentech	Case Name:	Hysys Distillation Unit.hsc
	Unit Set:	SI
	Date/Time:	Tue Feb 06 22:45:03 2018

WorkbocCase (Main)

Material Streams Fluid Pkg: All

Name		Kero_SS_Prod	Off Gas	Naptha	Diesel	AGO
Liquid Fraction		1.000	1.112e-005	1.000	1.000	1.000
Temperature	(C)	206	68.34	68.33	260.0	272.5
Pressure	(kPa)	183.8	239.2	239.2	291.0	296.1
Mass Density	(kg/m3)	637.2	4.755	631.3	651.9	660.5
Std Ideal Liq Mass Density	(kg/m3)	787.1	584.8	686.1	824.2	867.0
Std Ideal Liq Vol Flow	(m3/h)	7.000	2.409	33.000	20.00	2.000
Actual Liquid Flow	(m3/s)	2.403e-003	1.490e-005	9.877e-003	7.024e-003	7.292e-004
Actual Gas Flow	(ACT_m3/h)	Crude Oil Distillatio	Crude Oil Distillatio	Crude Oil Distillatio	Crude Oil Distillatio	Crude Oil Distillatio
Mass Flow	(kg/h)	5512	1409	2.245e+004	1.646e+004	1734
Molar Flow	(kgmole/h)	40.69	26.07	250.0	91.11	6.931
Molecular Weight		135.4	54.05	69.77	180.9	250.2
Viscosity	(cP)	0.2146	8.556e-003	0.2642	0.2420	0.1951
Name		**Residue**	**main steam**	**PA_2_Draw**	**PA_2_Return**	**Over Head Vapor**
Liquid Fraction		1.000	0.0000	1.000	1.000	0.000
Temperature	(C)	302.1	232.2	234.2	180	157
Pressure	(kPa)	306.2	1001	291.0	191.0	150
Mass Density	(kg/m3)	756.4	4.927	614.9	662.4	6.637
Std Ideal Liq Mass Density	(kg/m3)	963.5	998.0	814.5	814.5	689.2
Std Ideal Liq Vol Flow	(m3/h)	53.00	0.7515	34.45	34.45	90.83
Actual Liquid Flow	(m3/s)	1.870e-002	Crude Oil Distillatio	1.268e-002	1.177e-002	Crude Oil Distillation
Actual Gas Flow	(ACT_m3/h)	Crude Oil Distillatio	152.2	Crude Oil Distillatio	Crude Oil Distillatio	9431
Mass Flow	(kg/h)	5.106e+004	750.0	2.806e+004	2.806e+004	6.260e+004
Molar Flow	(kgmole/h)	153.3	41.63	164.5	164.5	782.6
Molecular Weight		333.0	18.02	170.5	170.5	79.99
Viscosity	(cP)	0.3464	1.725e-002	0.1810	0.2736	8.613e-003
Name		**Crude Oil**	**Reflux @COL1**	**To Condenser @COL1**	**Main Steam @COL1**	**Residue @COL1**
Liquid Fraction		1.000	1.000	0.0000	0.0000	1.000
Temperature	(C)	37.78	68.33	150.0	232.2	302.1
Pressure	(kPa)	101.4	239.2	273.7	1101	306.2
Mass Density	(kg/m3)	825.5	631.3	6.637	4.927	756.4
Std Ideal Liq Mass Density	(kg/m3)	842.1	686.1	689.2	998.0	963.5
Std Ideal Liq Vol Flow	(m3/h)	117.0	54.01	90.83	0.7515	53.00
Actual Liquid Flow	(m3/s)	3.315e-002	1.631e-002	Crude Oil Distillatio	Crude Oil Distillatio	1.870e-002
Actual Gas Flow	(ACT_m3/h)	Crude Oil Distillatio	Crude Oil Distillatio	9431	152.2	Crude Oil Distillatio
Mass Flow	(kg/h)	9.852e+004	3.706e+004	6.260e+004	750.0	5.106e+004
Molar Flow	(kgmole/h)	561.3	412.8	782.6	41.63	153.3
Molecular Weight		175.5	69.77	79.98	18.02	333.0
Viscosity	(cP)	5.290	0.2642	8.613e-003	1.725e-002	0.3464
Name		**Crude @COL1**	**Off Gas @COL1**	**Waste Water @COL1**	**Naptha @COL1**	**Diesel Steam @COL1**
Liquid Fraction		0.2761	0.0000	1.000	1.000	0.0000
Temperature	(C)	325.0	68.33	68.33	68.33	232.2
Pressure	(kPa)	344.7	239.2	239.2	239.2	1101
Mass Density	(kg/m3)	17.26	4.755	973.9	631.3	4.927
Std Ideal Liq Mass Density	(kg/m3)	842.3	584.8	998.0	686.1	998.0
Std Ideal Liq Vol Flow	(m3/h)	117.2	2.409	1.691	32.72	0.6363
Actual Liquid Flow	(m3/s)	1.907e-002	Crude Oil Distillatio	4.814e-004	9.877e-003	Crude Oil Distillatio
Actual Gas Flow	(ACT_m3/h)	Crude Oil Distillatio	296.3	Crude Oil Distillatio	Crude Oil Distillatio	126.9
Mass Flow	(kg/h)	9.872e+004	1409	1688	2.245e+004	635.0
Molar Flow	(kgmole/h)	572.4	26.07	93.69	250.0	35.25
Molecular Weight		172.5	54.05	18.02	69.77	18.02
Viscosity	(cP)	Crude Oil Distillatio	8.556e-003	0.4096	0.2642	1.725e-002

aspentech	Case Name:	Hysys .Distillation Unit.hsc
	Unit Set:	SI
	Date/Time:	Tue Feb 06 22:45:03 2018

WorkbocCase (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		Diesel @COL1	Diesel_SS_Draw @CC	Diesel_SS_Return @C	AGO @COL1	AGO Steam @COL1
Liquid Fraction		1.000	1.000	0.0000	1.000	0.0000
Temperature	(C)	217.2	244.2	234.5	272.5	232.2
Pressure	(kPa)	291.0	291.0	291.0	298.1	1101
Mass Density	(kg/m3)	651.9	614.9	5.801	660.5	4.927
Std Ideal Liq Mass Density	(kg/m3)	824.2	814.5	803.7	867.0	998.0
Std Ideal Liq Vol Flow	(m3/h)	20.00	26.47	7.109	2.000	9.090e-002
Actual Liquid Flow	(m3/s)	7.024e-003	9.741e-003	Crude Oil Distillatio	7.292e-004	Crude Oil Distillatio
Actual Gas Flow	(ACT_m3/h	Crude Oil Distillatio	Crude Oil Distillatio	985.0	Crude Oil Distillatio	18.41
Mass Flow	(kg/h)	1.648e+004	2.156e+004	5713	1734	90.72
Molar Flow	(kgmole/h	91.11	126.4	70.58	6.931	5.036
Molecular Weight		180.9	170.5	80.95	250.2	18.02
Viscosity	(cP)	0.2420	0.1810	1.135e-002	0.1951	1.725e-002
Name		**AGO_SS_Draw @COL**	**AGO_SS_Return @CC**	**PA_2_Draw @COL1**	**PA_2_Return @COL1**	**Kero_SS_Draw @COL**
Liquid Fraction		1.000	0.0000	1.000	1.000	1.000
Temperature	(C)	294.8	287.5	244.2	190.6	204.5
Pressure	(kPa)	298.1	298.1	291.0	291.0	283.8
Mass Density	(kg/m3)	625.5	4.687	614.9	662.4	610.5
Std Ideal Liq Mass Density	(kg/m3)	856.1	842.3	814.5	814.5	782.2
Std Ideal Liq Vol Flow	(m3/h)	2.550	0.6412	34.45	34.45	8.657
Actual Liquid Flow	(m3/s)	9.696e-004	Crude Oil Distillatio	1.268e-002	1.177e-002	3.081e-003
Actual Gas Flow	(ACT_m3/h	Crude Oil Distillatio	115.2	Crude Oil Distillatio	Crude Oil Distillatio	Crude Oil Distillatio
Mass Flow	(kg/h)	2183	540.1	2.806e+004	2.806e+004	6771
Molar Flow	(kgmole/h	9.433	7.538	164.5	164.5	50.81
Molecular Weight		231.5	71.65	170.5	170.5	133.3
Viscosity	(cP)	0.1498	1.440e-002	0.1810	0.2736	0.1778
Name		**Kero_SS_Return @CC**	**Kero_SS_Prod @COL**	**Kero Steam @COL1**		
Liquid Fraction		0.0000	1.000	0.0000		
Temperature	(C)	194.2	183.1	232.2		
Pressure	(kPa)	283.8	283.8	1101		
Mass Density	(kg/m3)	6.051	637.2	4.927		
Std Ideal Liq Mass Density	(kg/m3)	778.5	787.3	998.0		
Std Ideal Liq Vol Flow	(m3/h)	1.793	7.000	0.1364		
Actual Liquid Flow	(m3/s)	Crude Oil Distillatio	2.403e-003	Crude Oil Distillatio		
Actual Gas Flow	(ACT_m3/h	230.7	Crude Oil Distillatio	27.62		
Mass Flow	(kg/h)	1396	5512	136.1		
Molar Flow	(kgmole/h	17.67	40.69	7.554		
Molecular Weight		79.00	135.4	18.02		
Viscosity	(cP)	1.007e-002	0.2146	1.725e-002		