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Modelling and Economic Analysis of a Crude Oil Distillation System Using Aspen HYSYS

Abdulwahab Giwa[a,*], Oluwatodimu Adekunle Asiwaju[b], John Olusojiowolabi[c], Idowu Iyabo Olateju[d]

[a,b,c,d]Chemical and Petroleum Engineering Department, College of Engineering, Afe Babalola University, Ado-Ekiti, Ekiti State, Nigeria

[a]agiwa@abuad.edu.ng, [b]todimuasiwaju@gmail.com,[c]owolabijohn@abuad.edu.ng, [d]iolateju@abuad.edu.ng

*Corresponding author

Abstract

This work was focused on applying Aspen HYSYS to model, simulate and economically analyse a system for distillation of crude oil. The chemical components involved in the model development were water, methane, ethane, propane, i-butane and n-butane and some hypothetical ones that were added using light ends and TBP data with liquid volume basis before adding an output blend. Peng-Robinson was chosen as the fluid package. In the simulation environment, the crude was heated from 450 – 650 °F in a furnace before being fed into the distillation column, which was set up using Refluxed Absorber Column Sub-Flowsheet. The column had 29 stages and was also operated with a partial condenser. The outlet streams of the column were off gas, wastewater, naphtha, kerosene, diesel, AGO and residue. The developed model was simulated to convergence. The good convergence obtained from the simulation carried out on the developed Aspen HYSYS model of the atmospheric distillation unit showed that it (Aspen HYSYS) was able to handle the simulation of this process successfully. The Aspen Process Economic Analyzer tool was used to carry out an economic analysis of the developed model, and this generated an investment analysis. Therefore, it has been shown that the developed Aspen HYSYS model of the crude distillation unit that has been developed in this research work can be used to represent, simulate and analyse the cost of an atmospheric distillation unit successfully for further studies such as the development and testing of a prototype of the system as a small-scale petroleum refinery plant.

Keywords: Modelling, simulation, Aspen HYSYS, crude distillation, prototype, economic analysis, petroleum refinery.

1. Introduction

Process modelling is the representation, in form of a model, of engineering processes using expressions derived from combining knowledge of mathematics and others such conservation laws, thermodynamics, transport phenomena and reaction kinetics. A process model is a set of equations (including the necessary input data to solve the equations) that allow the prediction behaviour of a chemical process system [1] through simulation. Modelling and simulation are important in engineering because the description of system behaviour by experimentation may not be feasible due to inaccessible inputs and outputs, possible danger involved in experiment, very high cost and other constraints [2].

According to a classification, there are steady-state modelling and dynamic modelling. The former is vital during process conceptualization, design and evaluation [3] while the latter describes the change in the behaviour of system properties over time [4]. The system being referred to may be a petroleum refinery where crude distillation takes place.

An industrial process involving refining petroleum also involve crude distillation and is concerned with converting crude oil into products that are more beneficial [5, 6].

Crude distillation is known as the first major separation process as well as the fundamental process in the refining of crude oil. The pieces of equipment involved make up the crude distillation units (CDU) [7]. Changes in the CDU have great impact on product yields and quality. Therefore, CDUs are recommended to be operated at optimal conditions from both technical and economic standpoints.Process simulators that can be used to study the system and estimate the conditions required for optimal operation of the process are available in the literature.

The information obtained from the literature showed that some researchers have worked on some related topics. For instance, Muhammad *et al.* [8] simulated the crude distillation unit at Kaduna Refining and Petrochemicals Company (KRPC) with the aid of Aspen HYSYS. After the simulation, atmospheric residue had the highest volumetric flow rate of 313.7 m^3/h while AGO had the lowest with 10.90 m^3/h. Naphtha, kerosene and diesel had volumetric flow rates of 129.8 m^3/h, 42.14 m^3/h and 164.5 m^3/h, respectively. Therefore, it was concluded that the column needed to be optimized in order to convert more of the atmospheric residue into premium products like kerosene, gasoline etc. Parthiban *et al.* [9] successfully modelled and dynamically simulated a crude fractionation column with three side strippers using Aspen HYSYS Dynamics. The products were within the desired range and the expected specifications were attained. The results indicated that a multivariable chemical process such as fractionation column operation undergoes a clear nonlinear behaviour and can be modelled and dynamically simulated to observe thousands of process variations. Safwan bin Taharim [10] aimed at developing a steady-state model for CDU based on fundamental modelling approach using Aspen Plus. The effects of feed flow rate, feed composition and steam flow rate on product compositions and tray temperatures were studied. The results matched available data and the model was proved accurate. Goncalves *et al.* [11], using Aspen HYSYS, applied dynamic simulation to the atmospheric distillation unit of a crude oil refinery. The dynamic model developed was successfully combined with a suitable control configuration, thereby allowing detail study of the transient behaviour for new stationary levels of operation when subject to significant changes in operating conditions. Patrascioiu and Jamali [12] used the UniSim Design simulator to simulate the crude distillation process. In order to carry out this simulation, an analysis of the PRO/II, HYSYS, and Aspen HYSYS simulators was carried out. An initial implementation of specific problems of petroleum liquid-vapour equilibrium was carried out. The crude distillation simulation developed had one problem, though the program did not contain side strippers. It was therefore suggested that any future work should rectify that error. Khim [7] used Aspen Plus Dynamics to carry out the dynamic simulation of a crude distillation unit. An initial mathematical model was formed for the theoretical stage of the column. Based on the partial differential equations obtained, the relationship between the input and output variables was studied. From the results of the steady state simulation, it was proven that the change in energy balance of the column led to the change in mass balance. The dynamic model showed all changes of process variables with respect to time until system stability was achieved. Yin [2] used Aspen Plus to generate a model of crude distillation unit. By solving the model equations, the effects of different operating conditions of petroleum refining towards the yield and the composition of petroleum products were determined. The feed flow rates were found to vary directly with the products flow rates. To validate the simulation, the results were compared with the literature data available, and they were proven to be accurate. Fu [13] presented a hybrid model of a crude distillation unit suitable for planning, scheduling and real-time optimization (RTO). It eliminated the discrepancies between models used in these decision processes. Hybrid model consists of volumetric and energy balances and partial least squares model for predicting product properties. Product TBP curves were predicted from feed TBP curve, operating conditions (flows, pump-around heat duties, furnace coil outlet temperatures). Simulated plant data and model testing were based on a rigorous distillation model, with 0.5% RMSE over a wide range of conditions. Associated properties (e.g. gravity, sulphur) were computed for each product based on its distillation curve. Model structure made it particularly amenable for development from plant data. High model accuracy and its linearity made it suitable for optimization of production plans or schedules. Mustafa *et al.* [14] carried out a simulation of a crude distillation unit with Aspen HYSYS and presented a new method for the design of the distillation tower. From the simulation results, the light and heavy keys were determined to calculate the minimum number of trays and general dimensions of the distillation column. The design results were obtained to be number of trays, 52, diameter, 2.191 m and height, 26.6 m. The work also laid emphasis on the various types of control as well as tried to increase the productivities of naphtha from 7.75% to 11.62% and kerosene from 3.87% to 5.8%. Gas chromatography was used to validate the model. Velmurugan and Nalinakshan [15] carried out an Aspen HYSYS based simulation and analysis of a crude distillation unit. The feed into the pre-flash was at 491.5 tons/hr and the number of theoretical stages was specified to be 13. Experimental curves of kerosene, light gas oil and true boiling point curve of atmospheric residue were taken into account. Optimization was also easily carried out with the aid of advanced process control tools to make the system/process operational and profitable in real life.

Despite all these researches carried out so far concerning crude distillation, there is still insufficient supply of some petroleum products, especially gasoline at some period of the year. This might be due to the fact that more insight is required into the science and technology of the process. In order to contribute to that effect, this research work has been carried out to develop and simulate a crude oil distillation unit with the aid of Aspen HYSYS. After the modelling and simulation, the economic analysis of the process was also investigated for proper guidance.

2. Methodology

The methods used in this research were as outlined in the sections below.

2.1 Model Development and Simulation

Aspen HYSYSV8.4 [16] process simulator was used to develop the model of the crude distillation process. The light-end chemical components of the process were selected from the databank of Aspen HYSYS as seen in Figure 1. Also, Peng-Robinson was chosen as the property package (Figure 2).

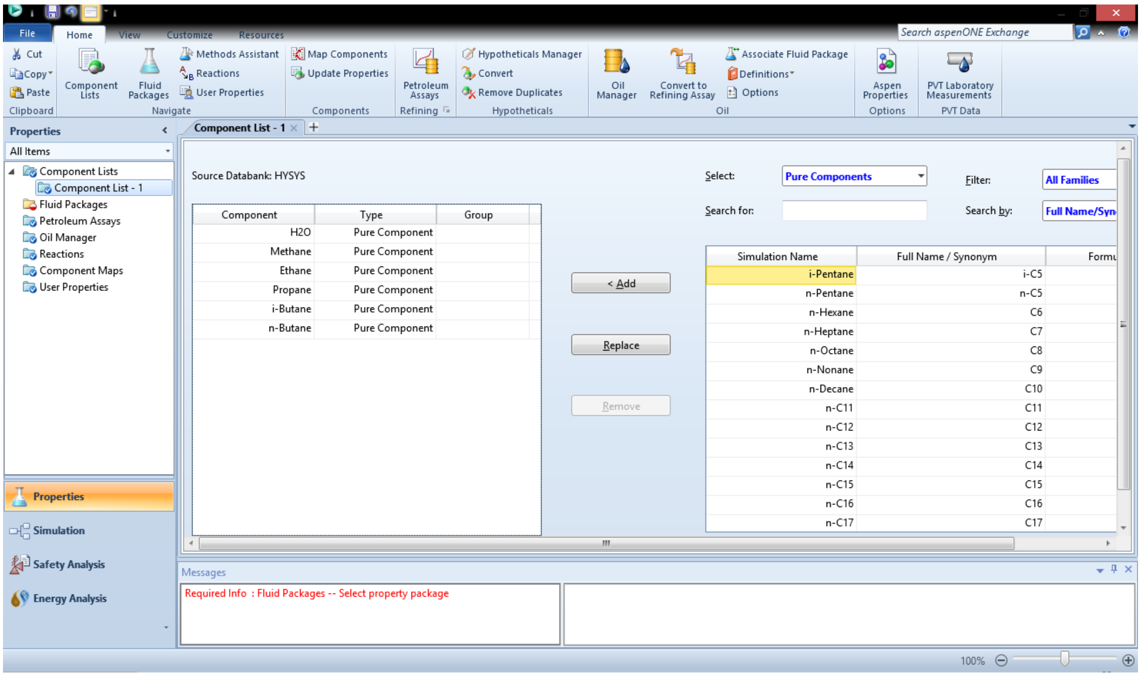


Figure 1: Adding chemical components

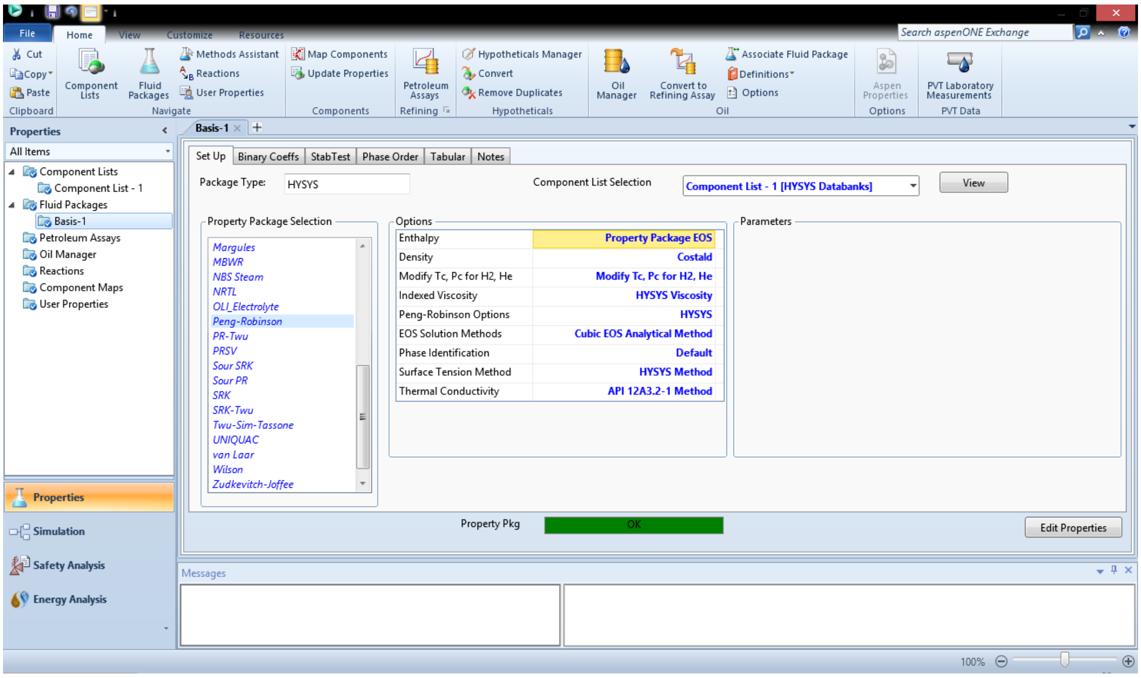


Figure 2: Selecting fluid package

After the component list and fluid package were completed, the oil manager environment where crude oil parameters and assay were inputted to the system was entered. In the Input Assay menu, the Assay Definition was completed. Bulk properties were entered as standard density of 29.32 °API, light ends and distillation data were typed in as shown in Figures 3 and 4 respectively. The assay was then calculated.

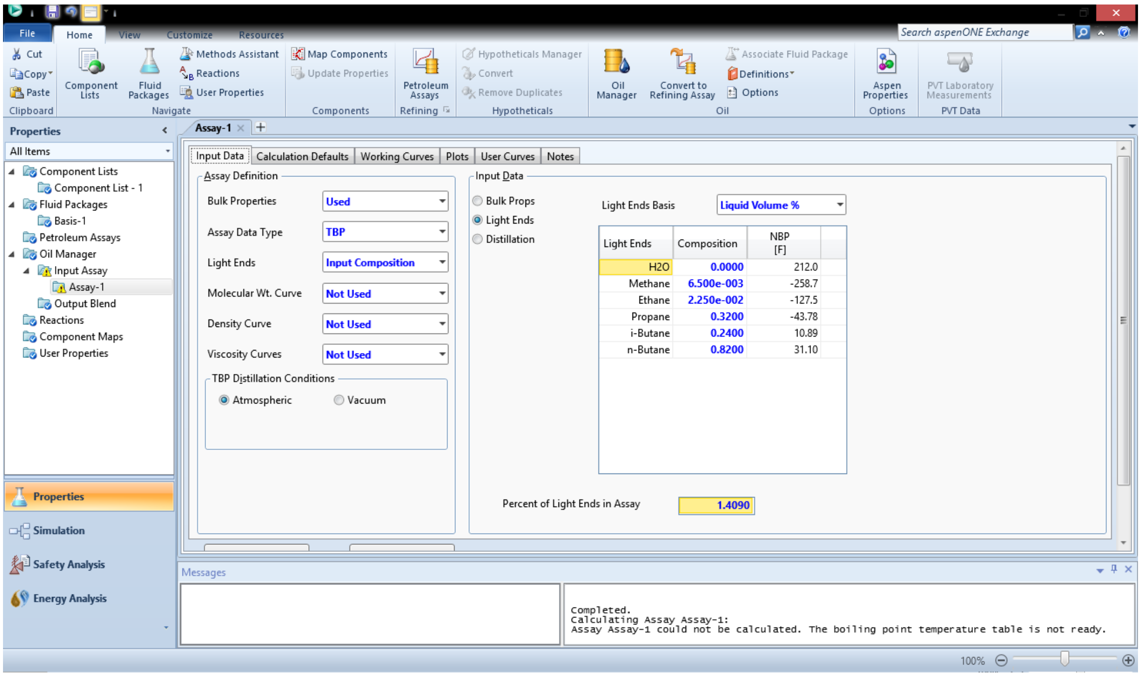


Figure 3: Entering Light Ends data

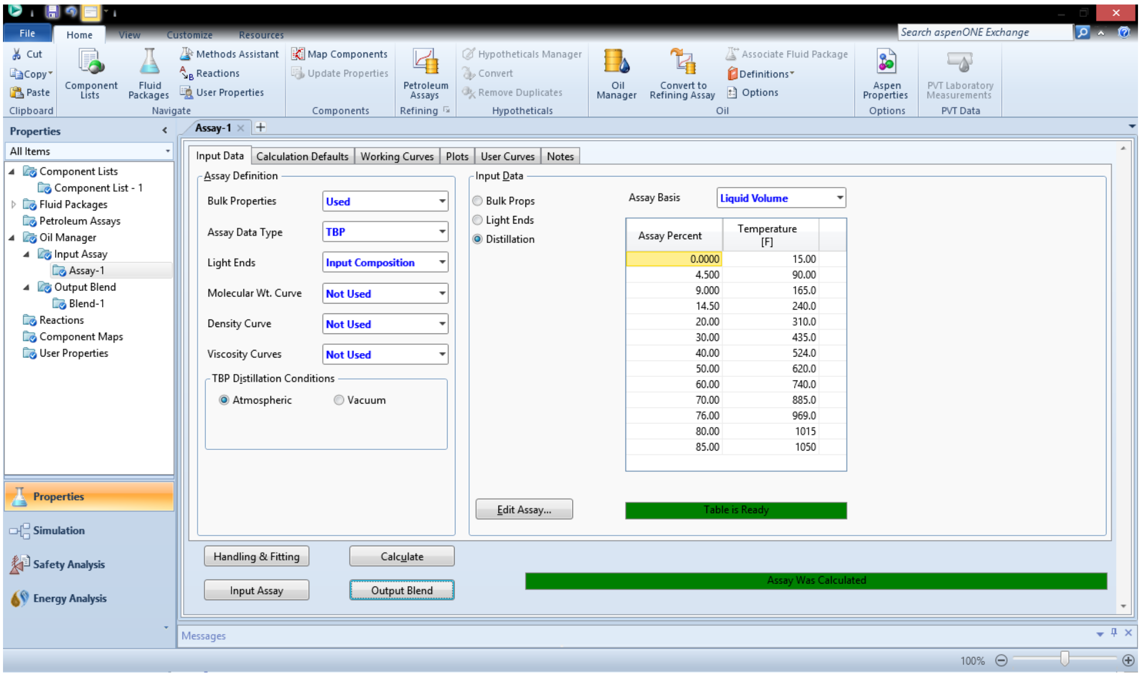


Figure 4: Inputting crude assay distillation parameters

After the oil input assay, the output blend was added and in 'cut ranges,' the user points option was selected and number of cuts was chosen to be 30. Oil was then installed and the stream was named 'Raw Crude.' The conditions for the raw crude were as given in Figure 5.

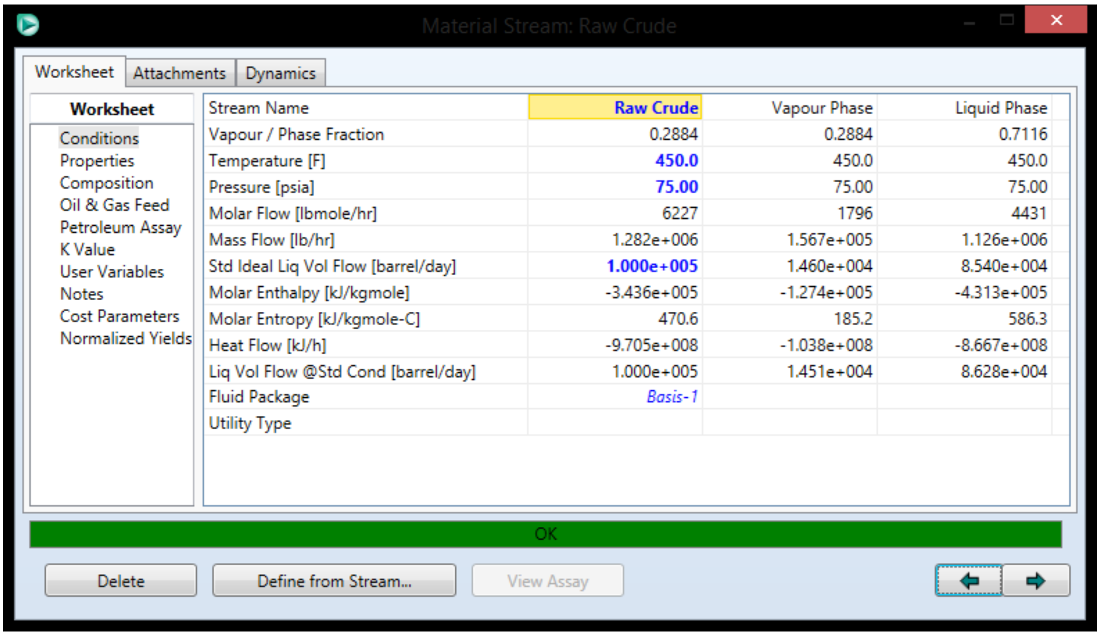


Figure 5: Inputting crude oil data

A heater (Figure 6) was then installed unto the flowsheet to heat the crude stream from a temperature of 450 °F to another temperature of 650 °F.

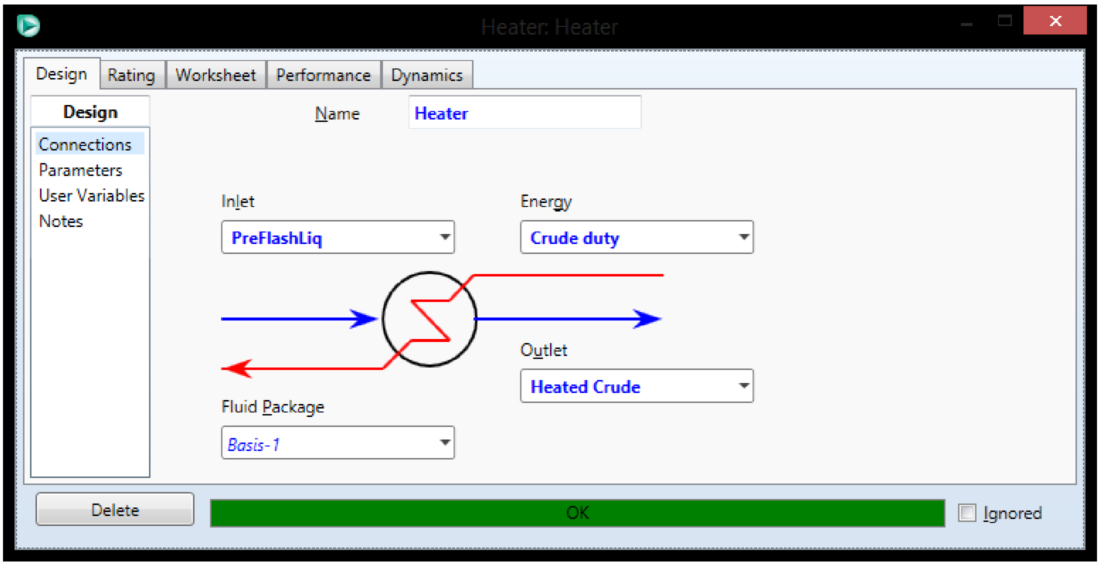


Figure 6: Crude oil heater installed

The steam streams for the atmospheric crude fractionator were installed and defined with the data given in Table 1.

Table 1: Steam streams data

Main Stream		
Temperature	375.0	F
Pressure	150.0	psia
Mass Flow	7500	lb/hr
Comp Mole Fraction	1.0000	
Diesel Stream		
Temperature	300.00	F
Pressure	50.00	psia
Mass Flow	3000.00	lb/hr
AGO Stream		
Temperature	300.00	F
Pressure	50.00	psia
Mass Flow	2500.00	lb/hr

Furthermore, the atmospheric crude fractionator having 29 stages (see Figure7) was added to the flowsheet using refluxed absorber column. The additional data (column pressure gradient) for the fractionator are given in Table 2.

Table 2: Column pressure gradient

Pressure drop of the condenser	9.00	psi
Pressure of the condenser	19.70	psia
Pressure of the bottom	32.70	psia

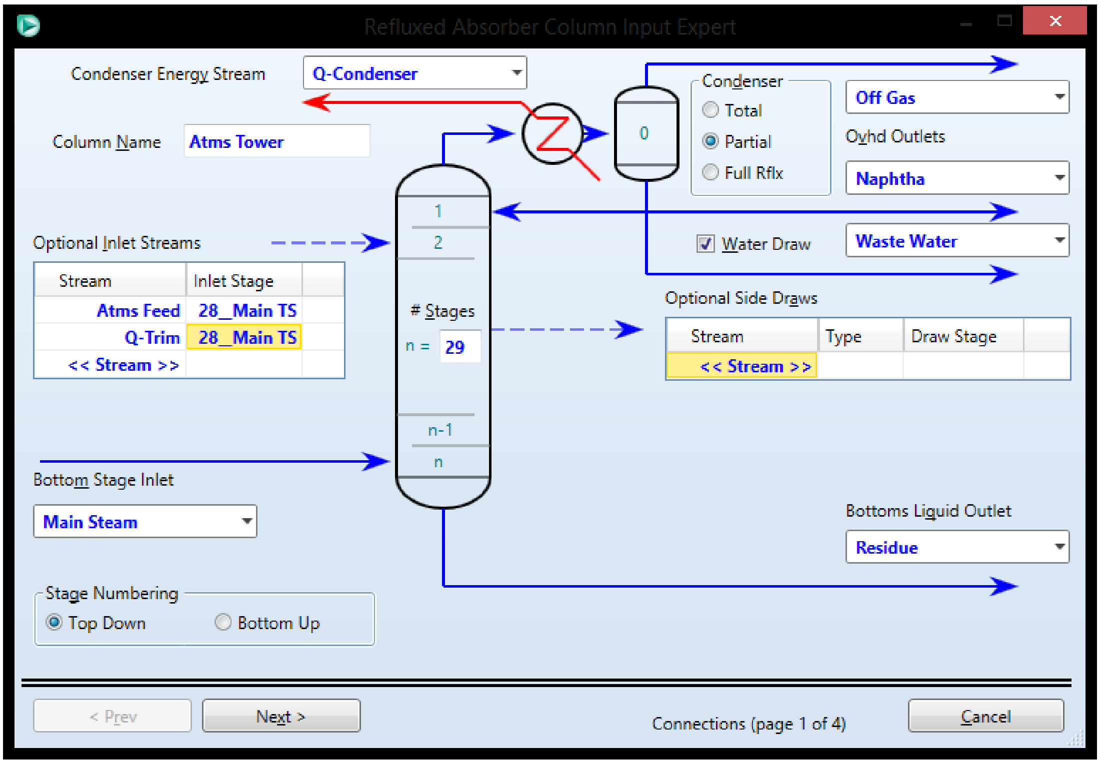


Figure 7: Entering crude fractionator data

Other given specifications were also added to the distillation column as shown in Figures 8 and 9 to obtain the final flowsheet of the crude oil fractionator given in Figure 10.

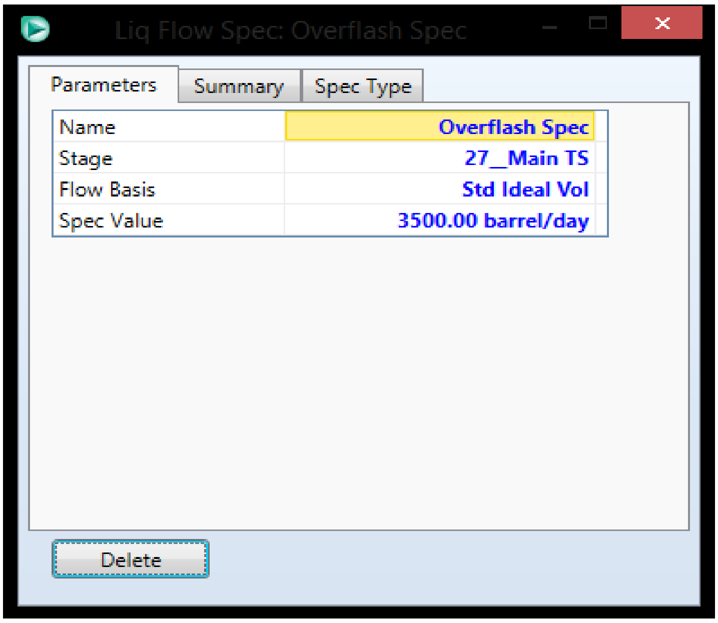


Figure 8: Adding column liquid flow specification

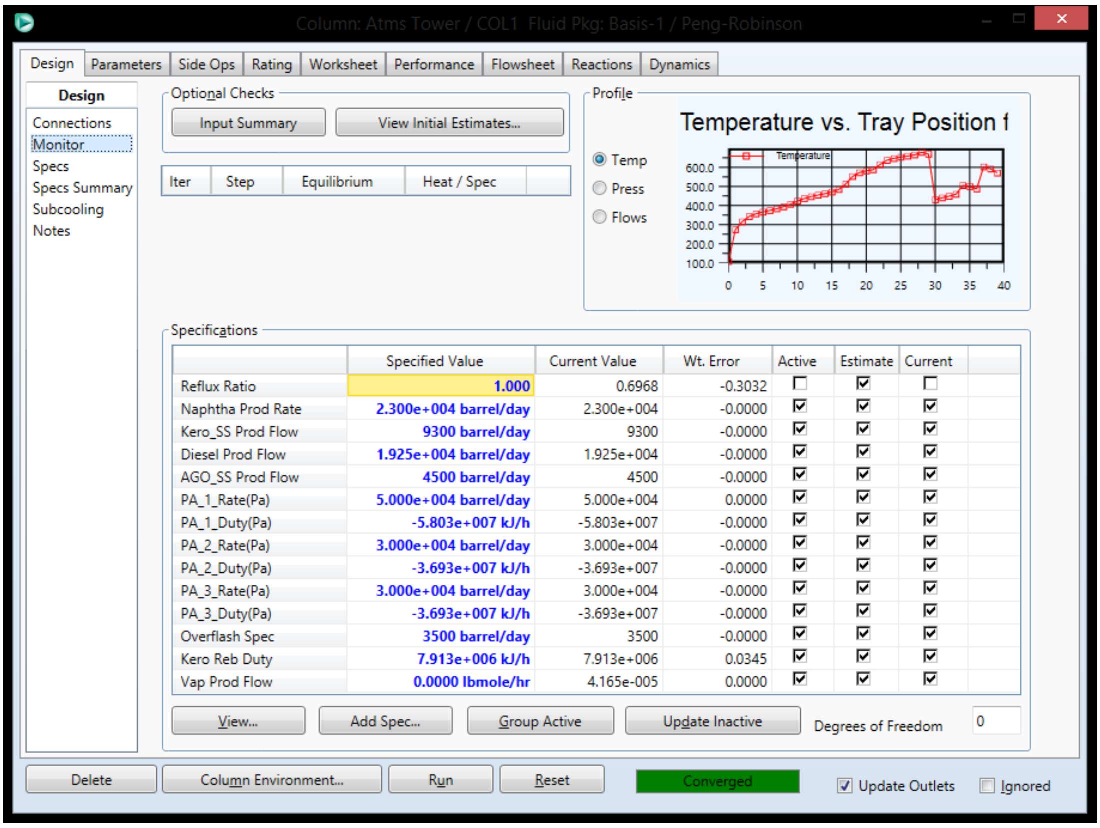


Figure 9: Column monitor

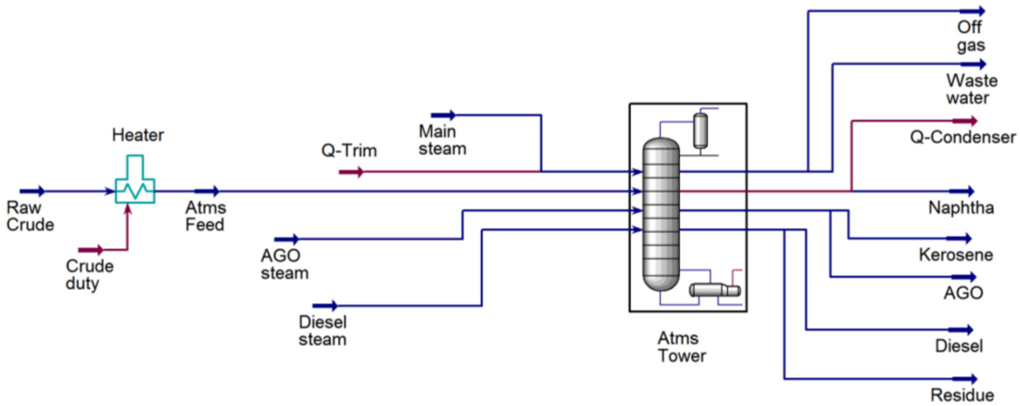


Figure 10: Atmospheric crude distillation flowsheet

2.2 Economic Analysis

After simulating the model of the atmospheric crude distillation process, its assessment was carried out on its economics using Aspen Process Economic Analyzer tool.

To accomplish that, the variables 'Cost Factor' including 'Cost Flow' were added in the 'Setup' menu of 'Workbook (Figure 11).' Then under 'Cost Parameters' in the worksheet of the 'Raw Crude' material stream, a cost factor was added. The cost factor was the crude oil price at the time, which was found to be about $57.41, see Figure 12.

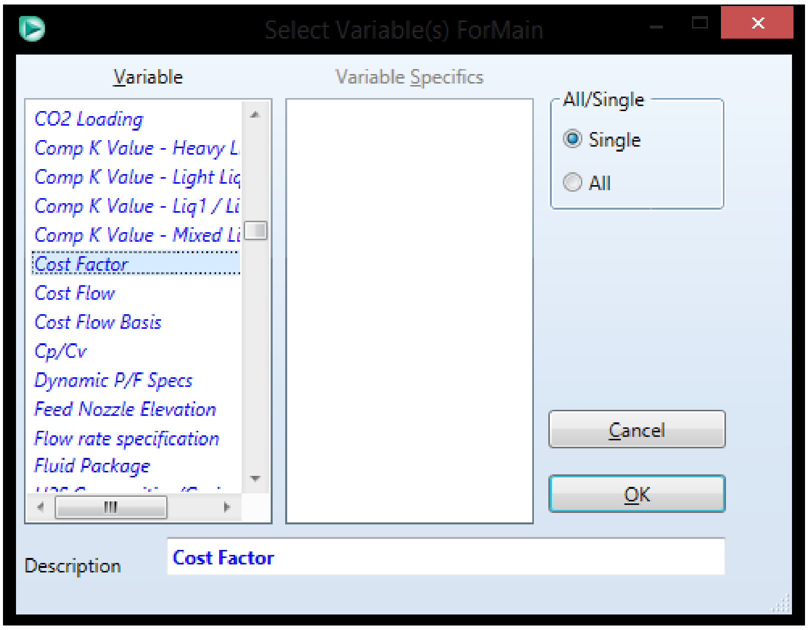


Figure 11: Adding cost factor variable

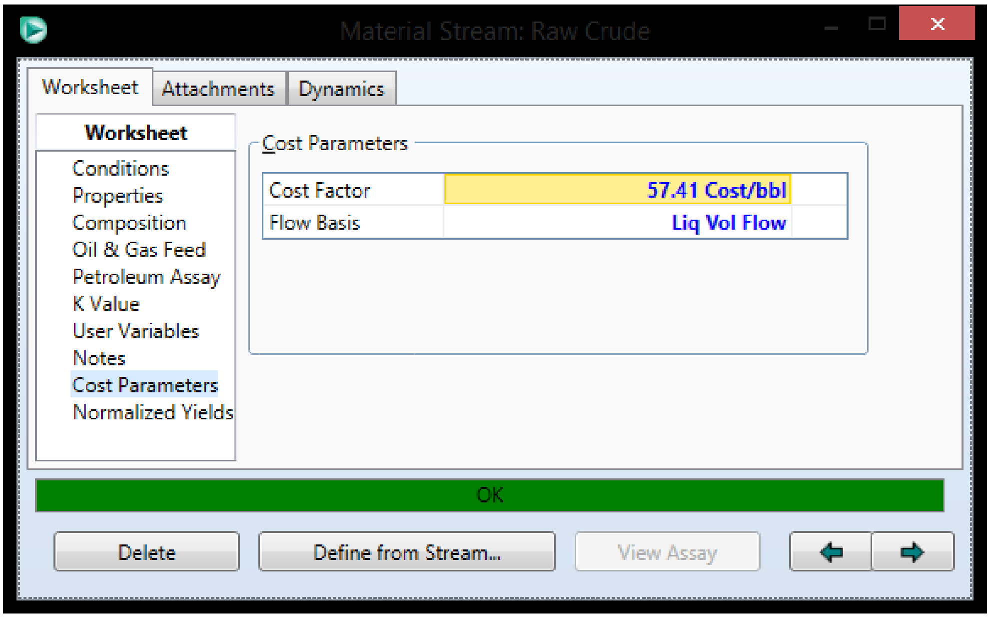


Figure 12: Cost parameters

In the ‘Economics’ tab, ‘Cost Options’ (Figure 13) form was opened to specify the costing template and other Aspen Process Economic Analyzer options, and currency symbols and conversion factors.

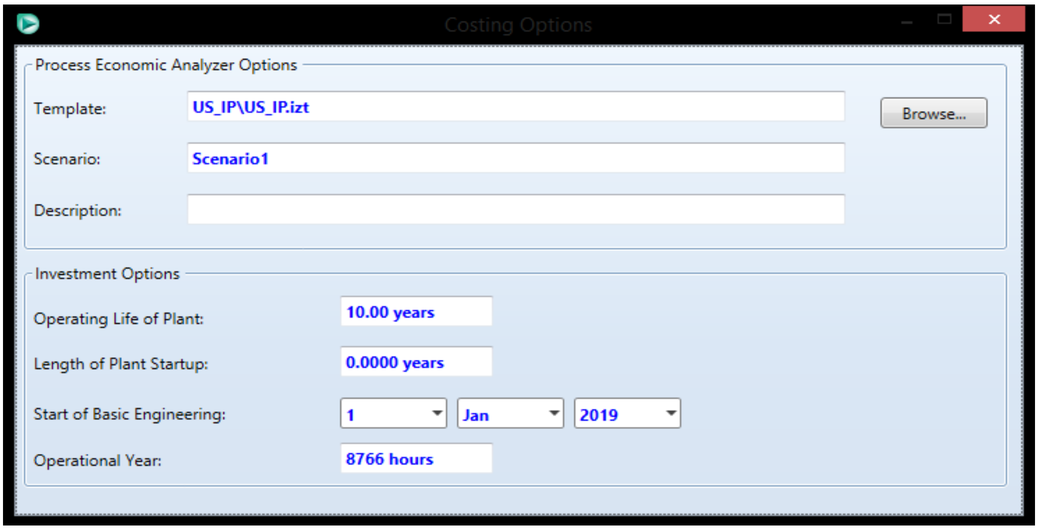


Figure 13: Adding cost options

To start the Aspen Process Economic Analyzer, the 'Economics Active' box was checked. 'Map' (Figure 14) was then clicked to view or modify the mapping of unit operation models to equipment models for economic evaluation. A pop-up box came up and the desired options were selected.

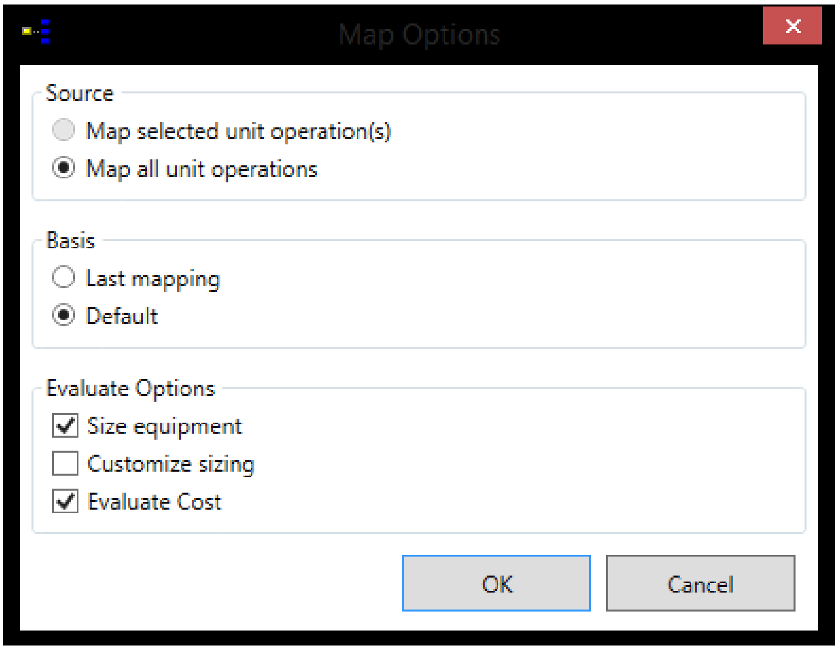


Figure 14: Mapping options

To perform sizing calculations using the Aspen Process Economic Analyzer, 'Size' was clicked. After that, 'View Equipment' was selected to display a summary of the results from the Economic Analyzer tool.

Finally, capital and operating costs were also evaluated with the aid of the Economic Analyzer of Aspen HYSYS process simulator.

3. Results and Discussion

The results given as the outputs of the research carried out in this work on the modelling, simulation and economic analysis of a crude distillation process for a petroleum refinery plant are as outlined and discussed thus.

3.1 Crude Characterization

Given in Figures 15 and 16 are the property plots obtained from the input assay and the output blend of the crude oil, respectively.

The results depicted in Figure 15 indicated that the true boiling point of the petroleum assay was increased as its volume was increasing. This was found to be in accordance with the principles of boiling because small volume is generally expected to boil faster than large volume. This was also observed to be an indication that the developed sample was a valid one.

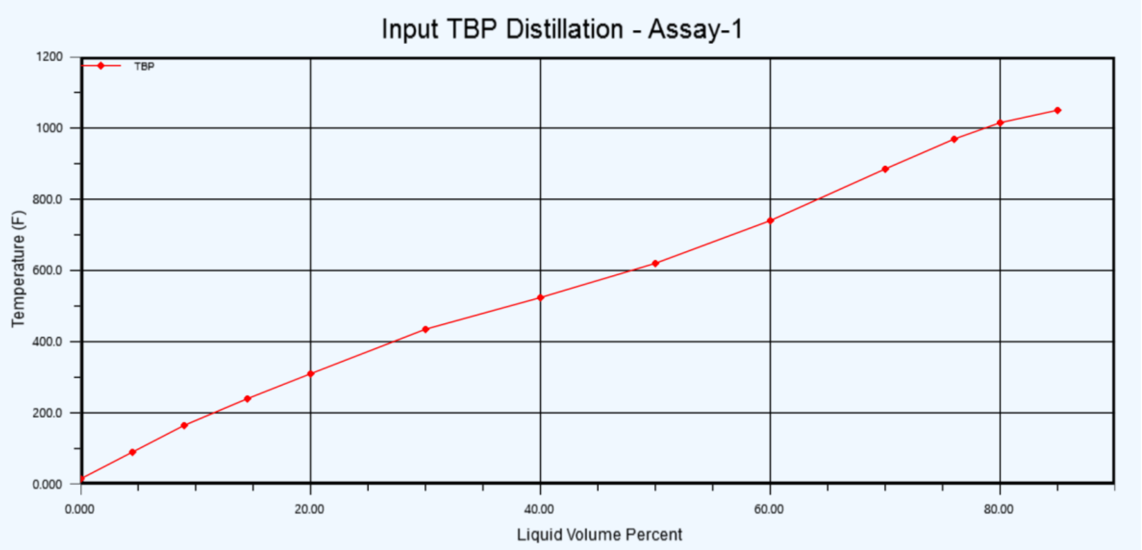


Figure15: TBP plot for input assay

Figure16: TBP plot for output blend

The trend given by the plot of true boiling points against the percentage liquid volume of the crude oil blend shown in Figure 16 indicated a similar relationship to that obtained in case of the assay. The reason for this case too was that large volumes of blend of crude oil boiled at higher temperature than small volumes.

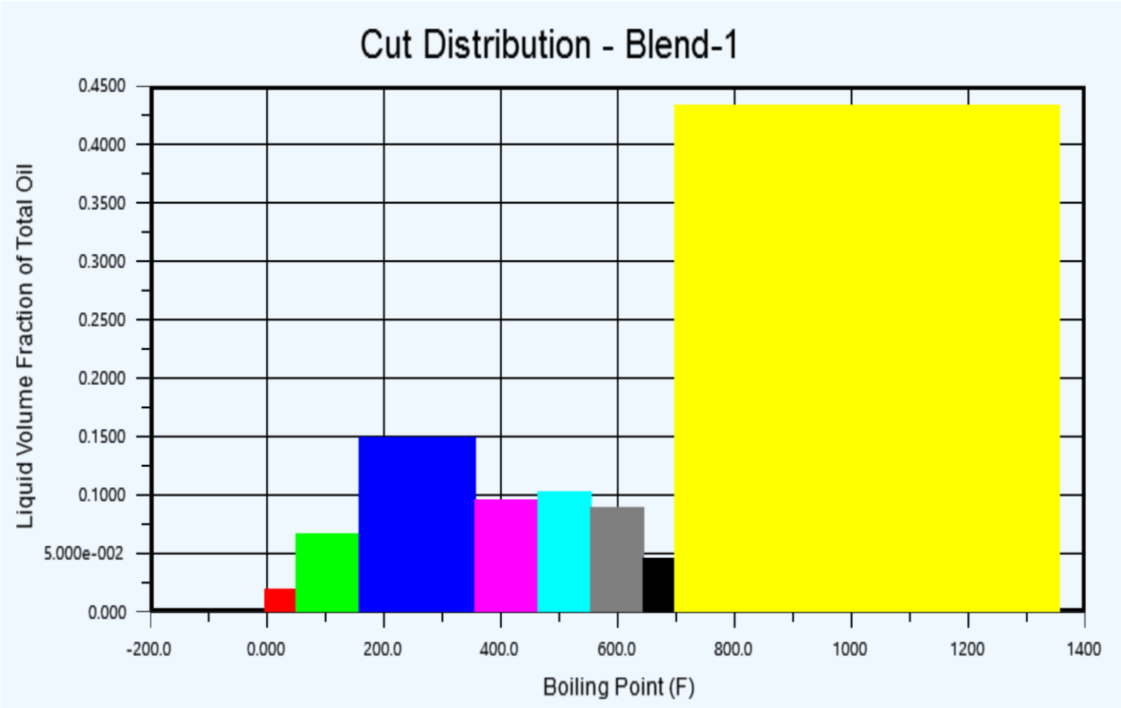


Figure17: Crude distribution curve

The crude distribution plot, given in Figure1717, shows the liquid volume fractions of the total oil at different boiling temperatures. In order of increasing boiling point, the fractions are: off gas, light straight-run naphtha, kerosene, light diesel, heavy diesel, AGO and residue. As it can be seen clearly from the curve, the residue had the highest quantity, at over 40% of total oil while refinery gases constitute the lowest fraction. Naphtha, from which gasoline is obtained, made up about 15% of total oil. The results shown in this figure were found to be supporting those given in Figure 16 in the sense that the fraction of the crude blend with the highest value had the highest boiling point while the one with the lowest fraction had the lowest boiling point.

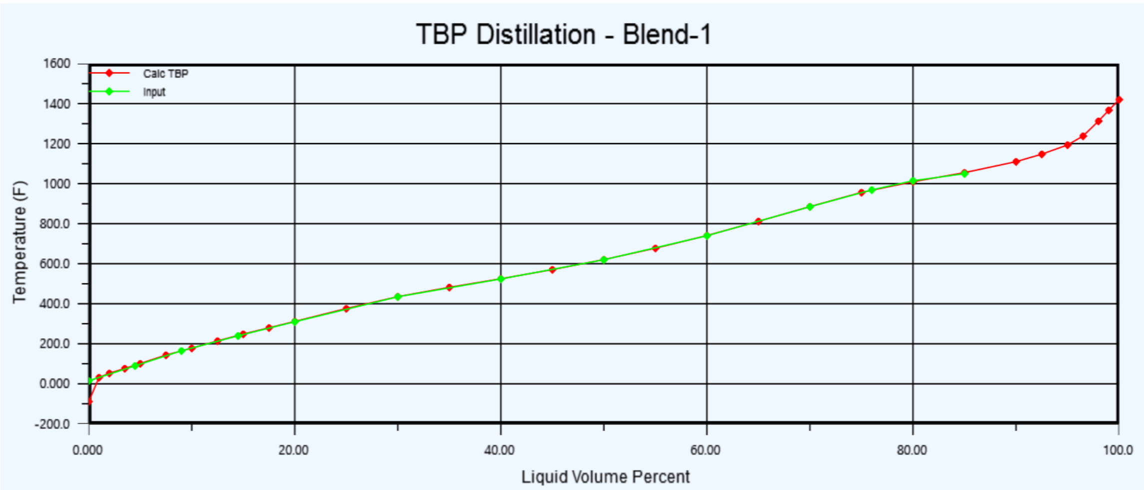


Figure 18: TBP and experimental curve comparison

Figure 18 shows the comparison between the experimental temperatures of the input assay and the calculated results. It can be seen very clearly from the figure that there are good agreements between both sets of data, which were found to be validating the data obtained from the developed process.

3.2 Material and Energy Balance

The material and energy flows obtained from the simulation of the crude oil distillation system model developed in this work are given in Table 3.

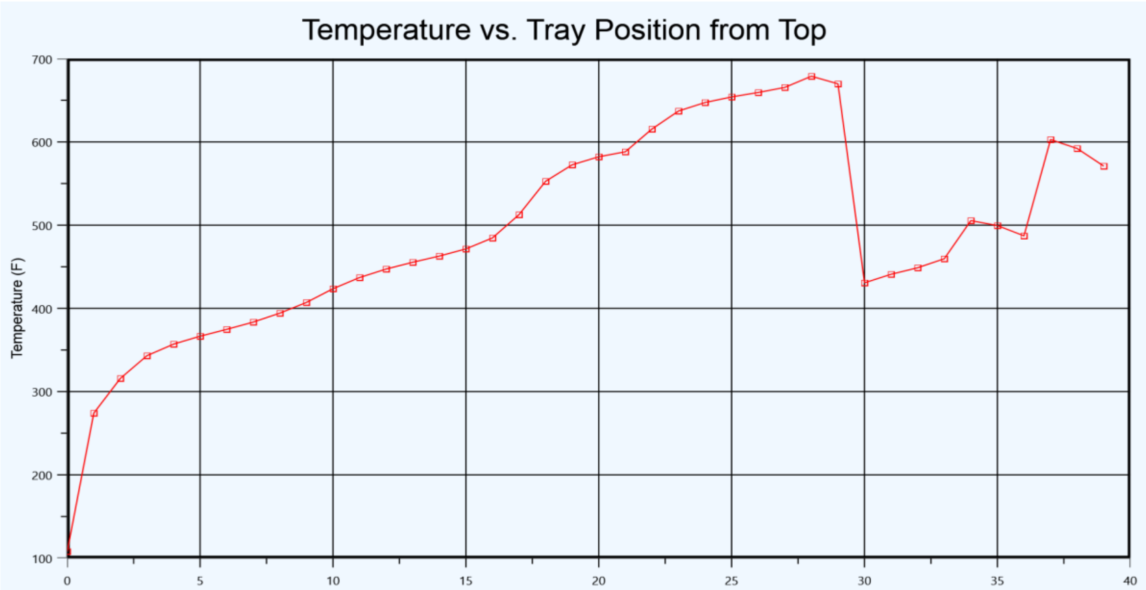
Table 3: Material and energy balances output values

Name	Vapour fraction	T [F]	P [psia]	Molar flow [lbmole/hr]	Mass flow [lb/hr]	Molar enthalpy [Btu/lbmole]
Atms Feed	0.6470	650.0	65.0	6227	1.282E+6	-1.154E+5
Main steam	1.0	375.0	150.0	416.3	7500	-1.018E+5
Diesel steam	1.0	300.0	50.0	166.5	3000	-1.022E+5
AGO steam	1.0	300.0	50.0	138.8	2500	-1.022E+5
Off gas	1.0	108.0	19.70	4.175E-5	2.158E-3	-5.151E+4
Naphtha	0.0	108.0	19.70	2824	2.474E+5	-8.221E+4
Waste water	0.0	108.0	19.70	701.4	1.264E+4	-1.225E+5
Kerosene	0.0	459.6	29.84	702.7	1.116E+5	-1.146E+5
Diesel	0.0	487.2	30.99	1117	2.429E+5	-1.534E+5
AGO	0.0	571.0	31.70	201.1	5.944E+4	-1.924E+5
Residue	0.0	669.8	32.70	1402	6.215E+5	-2.609E+5

The results shown in Table 3 also revealed the validity of the developed system because they (the results) were observed to be in line with the principles of distillation that is known in the literature. For instance, based on the observations made from the table, the known liquid products ranging from naphtha to residue were found not to have any vapour. This observation was also supported by the temperature of each of the products present in the system.

3.3 Column Profiles

Furthermore, the profiles of the crude distillation column were also considered in respect of some variables, viz: temperature, pressure and molar flow. The independent variable of the profile was taken to be the stage number of the column counted from the top down to the bottom.



Column stage
Figure 19: Temperature profile of the crude distillation column

Considering the temperature profile in **Error! Reference source not found.**19, the trend was observed not to be straightforward because the temperature was first increasing moving down the distillation column before a sharp drop was noticed at the point where kerosene was drawn. Thereafter, there was an increase in the temperature, though later dropped and continued increasing and decreasing until the reboiler. This non-particular nature of the temperature profile of the column is pointing to the complexity of the process, as expected.

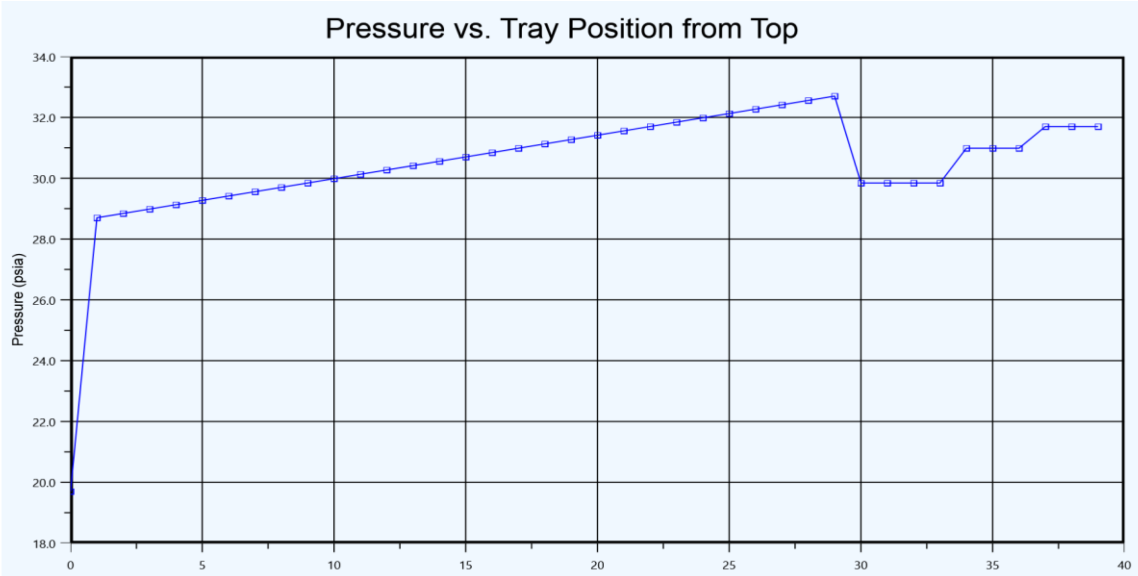


Figure 20: Pressure profile of the crude distillation column

The pressure profile of the column is given in Figure 20. The trend of the pressure profile shown in the Figure also indicated that the process was a complex one. This was due to the fact that the profile was found to be neither directly nor inversely proportional to the column stage.

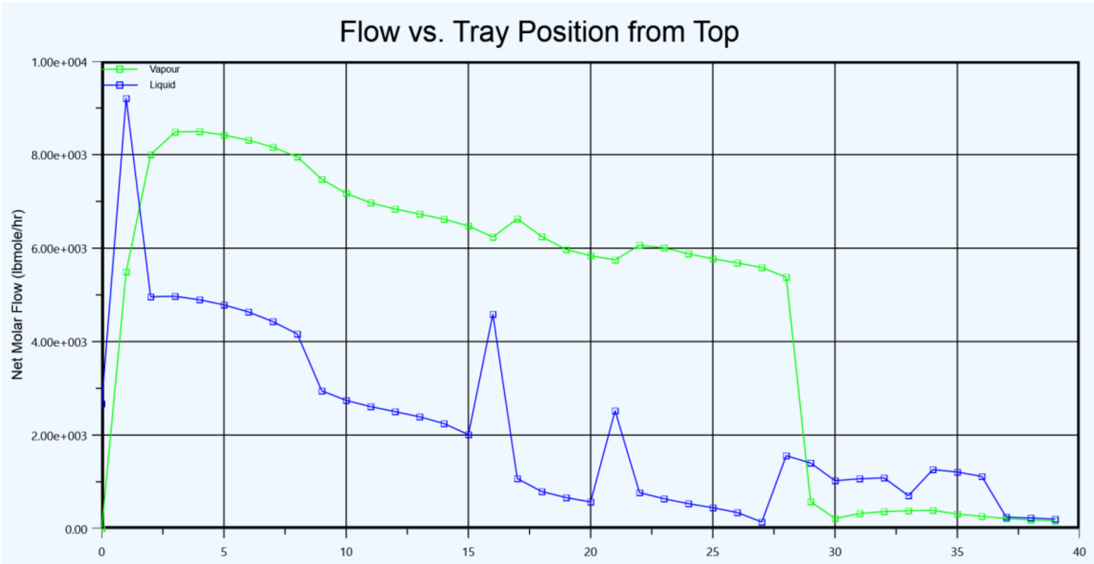


Figure 21: Molar flow profile of the crude distillation column

Given in Figure 21 are the molar profiles of both the vapour and the liquid component mixtures present in the crude distillation system developed in this work. According to the figure, it was observed that the liquid molar flow was significantly low in the condenser at the top of the column while there was no vapour flow there. The net liquid flow within the column was found to trend in a zigzag manner, in support of the complexities associated with the system. The net vapour molar flow of the mixture involved in the system was also found to have a profile that displayed a zigzag manner.

3.4 Economic Analysis

Table 4 gives the summary of the economic analysis of the crude distillation system modelled and simulated in this research work.

Table 4: Economic analysis results

Item	Unit	Value
Total project Capital Cost	Cost [USD]	1.43E+07
Total raw materials cost	Cost/period [USD/Year]	2.10E+09
Total products sales	Cost/period [USD/Year]	2.57E+09
Total operating labour and Maintenance Cost	Cost/period [USD/Year]	758040
Total utilities cost	Cost/period [USD/Year]	1.53E+06
Total operating cost	Cost/period [USD/Year]	2.27E+09
Operating labour cost	Cost/period [USD/Year]	657450
Maintenance cost	Cost/period [USD/Year]	100590
Operating charges	Cost/period [USD/Year]	164363
Plant overhead	Cost/period [USD/Year]	379020
Subtotal operating cost	Cost/period [USD/Year]	2.10E+09
Overhead cost	Cost [USD]	1.68E+08
Payback period	Year	5.23815

It was discovered from the results given in Table 4 that the process is profitable and economically viable because the total product sale was estimated to be greater than the operating cost. Also found out from the results was that the payback period (time to reach break-even point) was approximated to six (6) years by rounding up. This was another justification for the economic viability of the process.

4. Conclusion

The results obtained from simulating and modelling of a model of a crude distillation system having a column with 29 stages and operated at atmospheric pressure with the aid of Aspen HYSYS revealed that the product stream with the highest molar flow was the residue. Using the Aspen Process Economic Analyzer, an investment analysis of the model was obtained to be total project capital and total operating costs of 1.43E+07 USD and 2.27E+09 USD/Year, respectively. The results showed that the process was economically viable with a payback period of approximately 6 years. It is recommended that a prototype of the crude distillation unit be fabricated and tested for the development of a small-scale petroleum refinery plant.

Acknowledgment

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