MODIFICATION OF OUTLET STREAM OF THE ATMOSPHERIC DISTILLATION TO IMPROVE PRODUCTS FROM HEAVY CRUDE OIL USING ASPEN SIMULATIONS

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Abstract
Heavy crude oil distillation is a major challenge in the Nigeria's refinery of today. The refining process changes raw petroleum into about 2500 products. Some of the major products are distilled by simulation procedure in this work. These include gasoline, dual purpose kerosene, diesel and fuel oils. The simulation exercise starts with charging of heavy crude oil into a de-salter with the temperature condition of 250^{o}F. All other operating conditions such as the flow rate for the input crude oil was included in the worksheet. The distillation proceeded to a great extent and was controlled by the heavy crude oil that was fed into the preheater and the picked slate of oil based feeds. In order to obtain the targeted product, continuous refluxed was introduced into the simulation route. This was done to increase the output stream of the Atmospheric Distillation Unit. The product of the distillation obtained is in agreement with the general product cut expected from heavy crude oil. The success of these procedures which included the modified outlet stream are necessary for refinery's upgrade in Nigeria because of the energy saved.

Keywords: Refining, Distillation, Simulation, Heavy crude oil, Product

Introduction

Nigeria has witnessed a severe scarcity in the production 0f highway transportation fuels over the past three decades (Isa, et. al., 2013). The lack of productivity has been attributed to the inability of supply to meet the demand of the highway transportation fuel which has witnessed an unprecedented growth over the past decades (Isa, et. al., 2013). As much as 4 million litres in 2003, 5 million litres in 2004, 6 million litres in 2005, 3.42 million in 2006, 5 million litres in 2007, and 9 million litres in 2008 metric tons of highway transportation fuel were supplied from both local and imported fuel sources (Nwachukwu & Umunna, 2008). This shows that average amount of growth rate is 27.0% yet the amount of demand of transportation fuel fell short (Nwachukwu and Umunna, 2008). Due to this challenge, the content of Nigerian crude is of paramount interest and how to obtain sustainable fraction is a matter of urgency because of the current economic issues in the country. The elemental composition of Nigeria oils generally include carbon having 83-87%, hydrogen 10-14%, sulphur 0.05-6%, nitrogen 0.1-0.2%,oxygen -.05-2%, Ni < 120ppm vanadium<1200ppm (R0ussel and B0ulet, 1995). Traces of iron and copper are also present in the crude oil (Roussel and Boulet, 1995). Hydrogen to carbon ratios affects the physical properties of crude oil. As the hydrogen to carbon ratio decreases, the gravity and b0iling point of the hydrocarbon compounds increases (Fahim, et. al.,2009). The higher the hydrogen to carbon ratio of the feedstock, the higher its value to the refinery because less hydrogen is required. The composition of crude oil, on an elemental basis, falls within certain ranges regardless 0f its origin. Carbon and hydrogen c0ntents vary within narr0w ranges. F0r this reason, crude oil is not classified on the basis 0f carbon content. Despite their low concentrati0ns, impurities such as sulphur, nitrogen, oxygen and metals are undesirable because they cause concerns in the process ability of crude feedstock and because they affect the quality of the produced products (Fahim, et. al., 2009).

Multicomponent distillation involves separating complex mixtures by the volatilities the posses during heating to attain temperature at which each fractions will vaporize (Speight and Ozum, 2001). This method is based on the principle that different substances vaporize at different temperatures. For example, crude oil contains kerosene and naphtha, which are useful fractions (naphtha is made into petrol for cars, and kerosene is made into jet fuel) (Speight & Ozum, 2001). As the mixture of kerosene and naphtha vaporizes and condensed, the kerosene condenses at a higher temperature than the naphtha. As the temperature of mixture rises the kerosene condenses first, and the naphtha condenses later. Fractional distillation is separation of mixtures into its constituent fraction. Fractionation method is only economical if the process is operated on a large scale production and a continuous process (Wauquier, 1995). many plants design are batch distillati0n based because of the ease fabrication. This is because batch distillation allows for a high level of chemical purity and maximum flexibility.

Poisoning and corrosion are the most noticeable effects during refining. This research employs the use 0f ASPEN HYSYS for the numerical development of atmospheric distillation Unit (ADU) for the purpose of improving the cuts of gasoline and diesel from the distllati0n of petroleum cut such as naphtha, kerosene, diesel and AGO from heavy crude oil by increasing the outlet temperature of the atmospheric distillation unit (ADU). This will save the amount of energy needed in the VDU. It will also reduce waste, safe time and ultimately production cost (Hsu, & Robinson, 2017). Production, transportation, and refining of heavy crude oil present special challenges compared to light crude oil (Hsu, & Robinson, 2017). Generally, a diluent is added at regular distances in pipeline carrying heavy crude t0 facilitate its flow. Heavy oil is asphaltic and contains asphaltenes and resins (Gary, et.al., 2007). It is "heavy" (dense and viscous) due to the high ratio of aromatics and naphthenes to linear alkanes and high amounts of nitrogen, sulfur, oxygen and heavy metals (Hsu, & Robinson, 2017). High density oil has a higher percentage 0f compounds with over 60 carbon atoms and hence a high boiling point and molecular weight (Gary, et.al., 2007). F0r example, the viscosity of Venezuelas orinoco extra heavy crude oil lies in the range 1000-5000cP (1-5 pa.s), while Canadian extra heavy crude has a viscosity in the range 5000-10,000 cP (5-10 pa.s), about the same as molasses, and higher (approximately 100,00 cP or 100 pa.s for the most viscous commercially exploitable deposits) (Dusseault, 2001). Heavy crude oil is generally categorized in two ways. oil that have more than one percent of sulfur (high sulfur crude 0ils), with aromatics and asphaltenes and these are common in United State (California, Mexico), South America North America (Canada, Alberta, Saskatchewan), (Venezuela, Colombia & Ecuador) and the Middle East (Kuwait, Saudi Arabia) (Bunter, 2002). The oil containing less than one percent sulfur (low sulfur crude oils), with aromatics, naphthenes and resins, and these are common in Central Africa (Angola), Western Africa (Chad), and East Africa (Madagascar) (Clarke, 2010).

Process simulation is the use of a computer program to quantitatively model the characteristic equations of a process flow sheet by utilizing mass and energy balance, Equilibrium condition, rate correlations (reactions and heat-mass transfer processes). This will help to obtain an estimate of stream flow rates, compositions and properties, operating conditions, equipment sizes, steady state process simulation, rigorous petroleum simulation, petroleum/crude oil handling, data processing, best fit and optimum actualization. Modelling identifies an appropriate route in the distillation of heavy crude oil such as optimized energy requirement, temperature, and efficiency (Gary, et.al., 2007). ASPEN HYSYS is widely accepted and used for oil refining process. ASPEN HYSYS performs the oil distillation modeling by calculating details on plates. This calculation includes generating pseudo-constituent from the ASTM D86 data and obtain refining properties. ASPEN HYSYS is built to have an oil manager which orders the data for the pseudo-constituent separately (Begum, 2010).

In order to modify the output stream of the ADU using heavy crude oil as feed stock, the Aspen HYSYS software was employed to simulate atmospheric distillation. In this the Peng Robinson correlation was used to simulated the thermal effect by the iterative calculations of the thermo-physical and thermodynamic properties of the crude oil. Thus the petroleum products were produced by first pre-heating the heavy crude oil, desalting and distillation, modify reflux condition and hence increase the outlet temperature of the ADU. This is because the bulk fraction of heavy crude oil is the vacuum gas oil (VGO). Therefore Atmospheric Tower Bottom (ATB) product cut point will be increased. The cut point can be maintained at the threshold of the vacuum column diameter. An increase in the ATB cut point causes the vacuum unit feed to become heavier this gives a higher yield in ATB. By variation of the temperature of the vacuum heater outlet with be observed. The outlet temperature was raised using the setting in Aspen to about 700^{o}F for heavy Venezuelan oil which is similar to the heavy oil found in Ondo, Lagos, Ogun, and Edo State of Nigeria.

Methodology

Aspen HYSYS V8.6 was used to carry out the simulation of atmospheric distillation unit (ADU). On the user prop-1 form, the mass fraction option was chosen. The pure components values were set as zeroes to initiate the simulation of heavy oil composition. The density of crude oil was set. This procedure enabled the yield data. The crude oil assay was called; the sulphur distribution was added so that it can be tracked. To do the characterization calculation, crude oil blend was specified, and install into the flow sheet. At this point, a new blend of the crude was created and used as the feedstock in the simulation. When simulation was activated, a single stream called crude oil was presented in Figure 1. This stream was attached to two heat exchangers (to model the pre-heat before and after the Desalter) and a mixer to set an expected amount of water in the crude oil coming from the Desalter.

In order to improve the temperature and flow rate of the output stream of the Atmospheric Distillation Unint (ADU), the folowing are the conditions to be set on the operations. This value was chosen after several preliminary runs using some lower data from each feed and outlet. The most efficient value that yielded a modified output stream from the ADU are listed below:
Crude 0il Feed: 100^{0}F, 300 psig, 101,000 bpd
Preheat-1 0utlet: 260^{0}F, 294 psig
Desalter 0utlet: 260^{0}F, 294 psig, 500 bpd 0f water
Preheat-2 0utlet: 450^{0}F, 260psig

Crude oil stream was 0pened and the characteristic condition of the oil for heated crude oil which is the entry f0rm f0r next heater stream. The fl0w rate came fr0m 0il manager, pressure and temperature pr0vided were entered and the crude 0il calculati0n was c0mpleted and it t00k a new c0l0r in the fl0wsheet. Figure 1 presents the worksheet details.

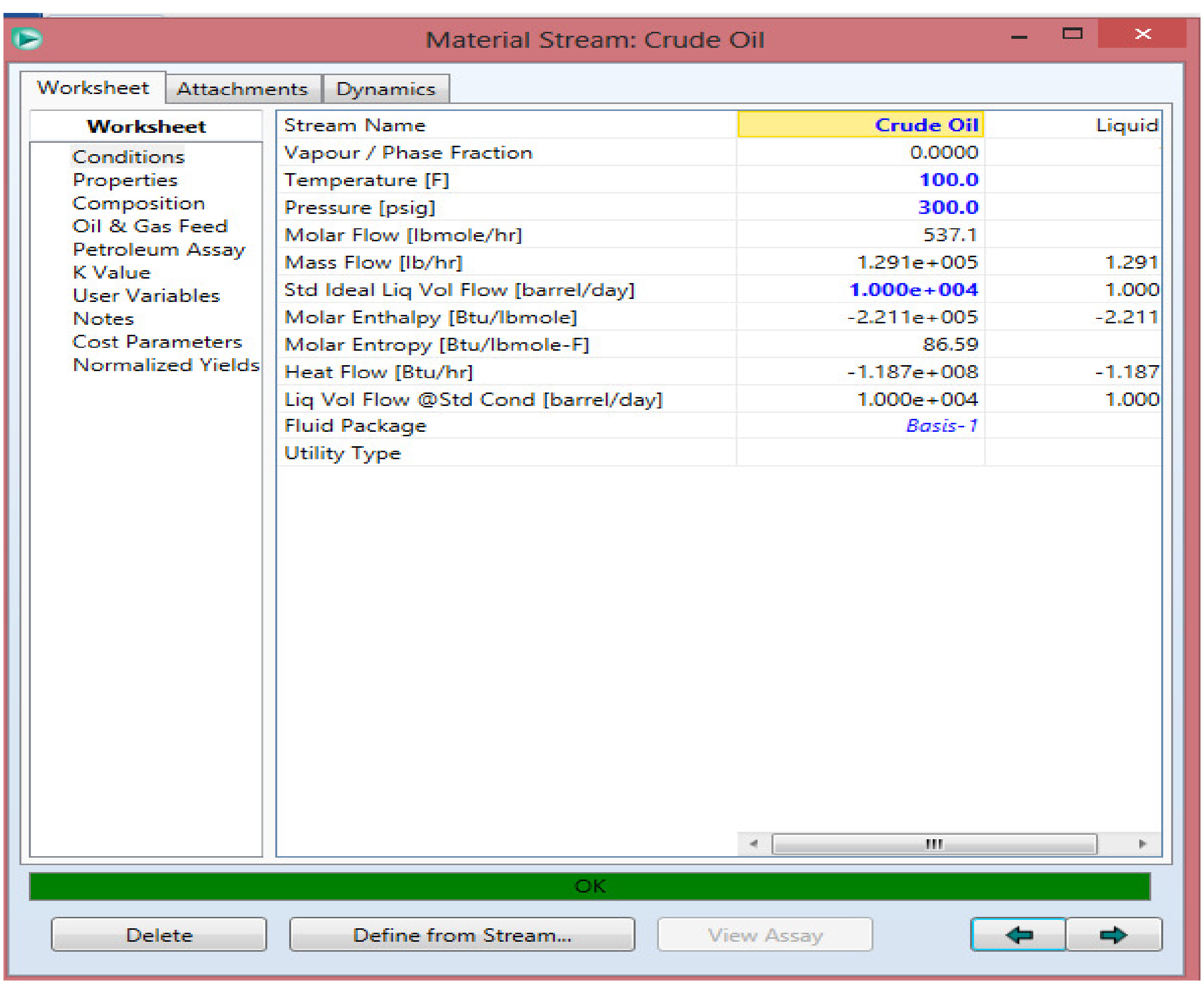

Figure 1: Crude oil streams

The ADU was included in the flow sheet, a refluxed absorber column was created on the fl0wsheet then the 50 trays were supplied.
In order to input the condenser energy stream, and set the total reflux, the c0lumn was activated 0n a wizard to obtain the bottom liquid outlets specified for an average of 70% efficiency. The basic inf0rmati0n f0r the main feeds and pr0ducts. The inf0rmati0n was filled in as sh0wn bel0w. The water draw b0x was checked.

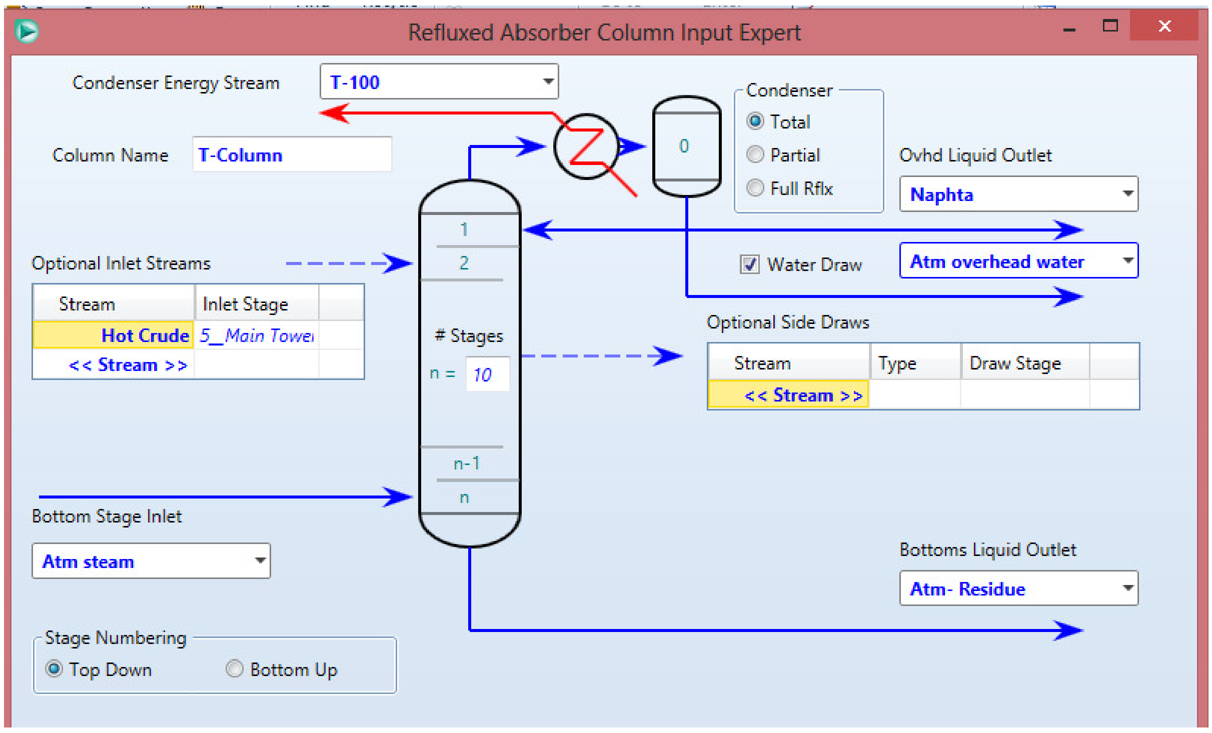

Figure 2: Refluxed Absorber Column

The basic pressure profile was set up in the column. The values of pressures were entered. Following this procedure, the estimate for the condenser temperature was set to 226.4°F for start. Next the atmospheric crude tower was included; the starting points were simulated from 250°F to 650°F for the top and bottom stages respectively.

The distillate rate estimate was set followed by the side strippers and pumparounds (with Draw from Tray 10, returned to Tray 7 for the kerosene). The first side operations screen was skipped since none of the side strippers is re-boiled. The basic information for the three side strippers were added so that at tray 10 the Kerosene was withdrawn, at tray 20, the diesel was withdrawn, at tray 30 the AGO was withdrawn. Because there was no vapour bypasses, side product flows were incorporated through the side strippers entered. The estimates of the flowrates out were entered in the bottom of the strippers and the 'NEXT' procedure was called. Next was to set the Specs on the pumparounds. The flowrate values was entered and the values associated with the heat exchanger duties. All the duty and temperature specs were returned T type was done and the next procedure was activated to obtain the improved temperature of 450 °F.

The pressure in the side strippers was set. A default value was used. Next procedure followed that the pressure drops across the pumparounds was set. Default values of zero was used then ended y activating done. The stage efficiencies were specified to model the stages as real trays. Under the parameters tab, 'efficiencies' was selected. Overall and User Specified Items (OUSI) was highlighted, and then efficiencies were applied.

Modification of the outlet stream of ADU
The steam of the streams was specified. This was achieved using the worksheet. The conditions were selected; the temperature, pressure and mass flowrate values were specified as listed in the procedure. The compositions was selected and set as a hundred percent of water. The normalize button was activated it was done. Produced petroleum are first desalted, introduced with steam to an atmospheric distillation column. The atmospheric

residue was then channeled into a vacuum distillation column operating at about 50 mmHg, where heavier products are retained.

Since it was desired to change the Heavy Gas Oil (HGO), Vacuum Gas Oil (VGO) and Vacuum Residue (VR), the distillation process was interrupted by increasing the temperature of the reflux and the residue had a temperature of 520-610^{o}F. The procedure adopted in this research included progressive reflux and repeated reflux which increased the temperature of the output stream of the ADU. This procedure serves as improved temperature for the residue which is channelled into the VDU. This has reduced the energy requirement for the Vacuum Distillation Unit (VDU). For this particular research the repeated reflux increased the temperature of the residue to about 850^{0}F.

Results and Discussion

Typical products from both columns and their boiling point ranges are Refinery gases (70-80^{o}F) , at 80 -90 ^{o}F the Liquid Petroleum Gases (LPG) was stilled 90-180 ^{o}F, the Light Straight Run (LSR) yielded at 100-180 ^{o}F, Next, the Heavy Straight Run (HSR) was obtained at 180 -380^{o}F, For the Dual purpose Kero (DPK) cut the temperature of yield was 380-400^{o}F. These results are in agreement with the work of Gary and Handwerk (2001).

Name	Hot Crude @COL1	steam @COL1	kero Steam @COL1	Diesel Steam @COL1	AGO Steam @COL1	Naphtha @COL1
Vapour	0.6855	1.0000	1.0000	1.0000	1.0000	0.0000
Temperature [C]	334.4	260.0	260.0	260.0	260.0	42.70
Pressure [kPa]	273.7	1136	1825	1825	1825	128.9
Molar Flow [kgmole/h]	2644	7340	917.4	917.4	917.4	1355
Mass Flow [kg/h]	5.949e+005	1.322e+005	1.653e+004	1.653e+004	1.653e+004	1.465e+005
Std Ideal Liq Vol Flow [m3/h]	672.4	132.5	16.56	16.56	16.56	200.1
Molar Enthalpy [kJ/kgmole]	-3.142e+005	-2.342e+005	-2.345e+005	-2.345e+005	-2.345e+005	-2.359e+005
Molar Entropy [kJ/kgmole-C]	747.8	173.0	168.7	168.7	168.7	128.2
Heat Flow [kJ/h]	-8.307e+008	-1.719e+009	-2.151e+008	-2.151e+008	-2.151e+008	-3.197e+008
Name	water Residue @COL1	Atm Residue @COL1	Kerosene @COL1	Diesel @COL1	AGO @COL1	
Vapour	0.0000	0.0000	0.0000	0.0000	0.0000	
Temperature [C]	42.70	278.4	172.6	215.2	226.4	
Pressure [kPa]	128.9	253.0	219.2	227.7	236.1	
Molar Flow [kgmole/h]	1.029e+004	584.0	237.9	223.6	72.56	
Mass Flow [kg/h]	1.854e+005	3.158e+005	4.851e+004	5.855e+004	2.249e+00	

Figure 3: Worksheet of the simulation

The basic typical properties of a naphtha product stream during crude simulation are presented in the worksheet in Figure 3.

Table 1: Naphtha product stream properties for the simulation

Properties	Base case
Vapour	0.00
Temperature [^{0}C]	42.68
Pressure [kpa]	128.9
M0lar flow [kgmole/hr]	1355
Mass flow [kg/hr]	1.465x10^{5}
Std ideal Liquid Vol. flow [m^{3}/hr]	200.1
M0lar Enthalpy [kJ/kgmole]	-2.359x10^{5}
M0lar Entropy [kJ/kgmole-C]	128.2
Heat flow [kJ/hr]	197

The Table 1 presents the values of temperature, pressure, molar flow, mass flow, volumetric fl0w, molar enthalpy, molar entropy and heat flow for Naphtha product stream for the simulation. In Table 1, a revamp up process stream must obtain for the heavy crude oil preheated and related with preparing substantial unrefined while keeping up unit dependability. Higher-viscous heavy crudes (temperature at 42^oC) diminish the crude charge pump (200.1m^3/hr) created head and can likewise expand exchanger fouling. Evading water driven points of confinement to accomplish a coveted unrefined enhance flow rate (1355kgmol/hr) and corresponding mass flow of 1.465x10^5 which can be costly. Therefore, a revamp must be considered with crude pressure pump.

The basic typical properties of a kerosene product stream during crude simulation are presented in Table 2.

Table 2: Kerosene product stream properties for the simulation

Base case	properties
Vapour	0.00
Temperature (^0C)	172.6
Pressure (kpa)	219.2
Molar flow [kgm0le/hr]	237.9
Mass flow [Ib/hr]	4.851x10^4
Std ideal Liquid Vol. flow [m^3/hr]	58.30
Molar Enthalpy [kJ/kgmole]	-3.751x10^5
Molar Entropy [kJ/kgmole-C]	411.8
Heat flow [kJ/hr]	-8.925x10^7

Table 2 presents the values of temperature, pressure, molar flow, mass flow, volumetric flow, molar enthalpy, molar entropy and heat flow for kerosene product stream for the simulation. Revamp process design is done in Table 2 to Table 5 for kerosene product stream. Particularly in Table 2 crude - unit process stream plans was centred around crude oil properties, to vaporize kerosene and further heating of the crude oil to distil heavier product later. Heavy oil is intrinsically difficult to vapourize on the grounds that there is less light materials in the bolster. The crude oil unit design must advance product from ADU to adjust general unit performance since it impacts crude and vacuum unit temperature of 172.6 oC. The connection amongst ADU and VDU cut focuses are complex and distillation progresses by the design adjustment on the temperature. With a lower temperature of the ADU product, there is less atmospheric of zero at the kerosene plate accessible to give pumparounds and content heated to initial crude pre-heated.

The basic typical properties of a diesel product stream during crude simulation are presented in Table 3.

Table 3: Diesel pr0oduct stream properties for the simulation

Properties	Base case
Vapour	0.00
Temperature [^0C]	215.2
Pressure [kpa]	227.7
Molar flow [kgmole/hr]	223.6
Mass flow [kg/hr]	5.855x10^4
Std ideal Liquid V0l. flow [m^3/hr]	67.83
Molar Enthalpy [kJ/kgmole]	-4.508x10^5
Molar Entropy [kJ/kgmole-C]	627.1
Heat flow [kJ/hr]	-1.008x10^8

Table 3 shows the values of temperature, pressure, molar flow, mass flow, volumetric flow, molar enthalpy, molar entropy and heat flow for Diesel product stream for the simulation.
Table 3 presents great diesel product initial condition at vapour zero, the diesel-AGO product fractionation requires several reflux rate (fluid - vapour proportion), the column section is at the 8-10 plate which emerge a good plate productivity due to the modification of the crude input. Most atmospheric column wider than 16ft in measurement will utilize

four plates or more. These modified breadth towers delivers low weir loadings (100gpm/in. weir) and the plate effectiveness was as low as 50%. Although a low reflux and plate efficiency significantly diminish diesel yield. Therefore the reflux was increased which significantly increases the plate efficiency. The basic typical properties of an atmospheric gas oil product stream during crude simulati0n are presented in Table 4.

Table 4: Atmospheric Gas 0il product stream properties for the simulation

Properties	Base case
Vapour	0.00
Temperature [^{0}C]	226.4
Pressure [kpa]	236.1
Molar flow [kgm0le/hr]	72.56
Mass flow [kg/hr]	$2.249 x 10^4$
Std ideal Liquid Vol. flow [m^3/hr]	25.40
Molar Enthalpy [kJ/kgm0le]	$-5.241 x 10^5$
Molar Entropy [kJ/kgm0le-C]	780.6
Heat flow [kJ/hr]	$-3.803 x 10^7$

Table 4 shows the values of temperature, pressure, molar flow, mass flow, volumetric flow, molar enthalpy, molar entropy and heat flow for AGO product stream for the modelling. Table 4 presents the initial condition for the start of the Atmospheric Gas Oil (AGO) pumparound builds crude oil preheat. The initial vapour for diesel was at zero. Further, in any case, if the ADU product is only 700^{4o}F, there is not enough vapour from the atmospheric distillation column flash zone to give adequate inner reflux in the diesel-AGO fractionation unit to permit heat and the removal from an AGO pumparound. This phenomenon is controlled by the mass flow rate of $2.249 x 10^4$ operating with an AGO pumparound was based on crude True Boiling Point (TBP) for improved distillation, adequate diesel-AGO fractionation, and ATB cut point target using the parameter for standard ideal liquid volume flow of 25.40m ^{3}/hr. The basic typical properties of an atmospheric residue product stream during crude simulation are presented in Table 4.

Table 5: Atmospheric Residue Product stream properties for the simulation

Properties	Base case
Vapour	0.00
Temperature [^{0}C]	278.4
Pressure [kpa]	253.0
Molar flow [kgm0le/hr]	584.0
Mass flow [kg/hr]	$3.158 x 10^5$
Std ideal Liquid Vol. flow [m^3/hr]	317.7
Molar Enthalpy [kJ/kgm0le]	$-8.603 x 10^5$
Molar Entropy [kJ/kgm0le-C]	1594
Heat flow [kJ/hr]	$-5.024 x 10^8$

Table 5 presents the values of temperature, pressure, molar flow, mass flow, volumetric flow, molar enthalpy, molar entropy and heat flow for Atm0ospheric Residue product stream for the simulation. Table 1 to 5 are the process simulation results. The simulated volumetric rate with corresponding temperature and pressure for Naphtha product are 200.1m^3/hr, 42.68^{0}C and 128.9kpa. The simulated volumetric rates with corresponding temperatures and pressures for Kerosene product, Diesel product, AGO product and ATM residue product were 58.30m^3/hr, 172.6^{0}C and 219.2kpa, 67.83m^3/hr, 215.2^{0}C and 227.7kpa, 25.40m^3/hr, 226.4^{0}C and 236.1kpa, and 317.7m^3/hr, 278.4^{0}C and 253.0kpa as shown in table 1 to 5 respectively. This shows that the higher the temperature and pressure, the higher the volumetric rate of the product. The vapour phase for all products is zero which indicates that they are completely in liquid phase.

Operating with an AGO pumparound was based on crude True Boiling Point (TBP) for improved distillation, adequate diesel-AGO fractionation, and ATB cut point target using the parameter for standard ideal liquid volume flow of 25.40m3/hr. The vacuum-unit configuration relies upon the HVGO product cut point target, vacuum heater dsign, unrefined vanadium dispersion, and other Nitti gritty hardware configuration issues. HVGO product cut.

Focuses are regularly under 975 ^{o}F. at the point when crudes with gravities under 24 ^{o}API are being prepared. A dry vacuum unit configuration utilizes no stream in the radiator and does not have a stripping area. Keeping up cut point is difficult even with an good design unit utilizing coil steam, yet a dry vacuum unit essentially do not work dependably at cut prominent than around 950 ^{o}F. When preparing Akure, Egbama and creek crude oil. A heater without coil steam must work at 760-770 ^{o}F to keep away from fast boiling from havy crudes. HVGO item TBP cut points> 1000 ^{o}F. Require a warmer outlet temperature of 795-800 ^{o}F, low glimmer zone oil fractional weight and great VTB stripping. Overwhelming unrefined builds add up to LVGO and HVGO pumparound obligation necessities since more VGOs are yielded. A two-item vacuum segment will have a high HVGO pumparound obligation at a generally low temperature of around 480 ^{o}F -540 ^{o}F. Expanding vacuum unit heat input requires more surface region and more HVGO pumparound obligation normally requires increasing the quantity of exchangers in arrangement in light of the fact that the log mean temperature contrast is so low. One refiner utilized six exchangers in arrangement in the hot preheater prepare. Exchanger organize configuration must address the expanding weight drop caused by extra exchangers.

This refiner can then again incorporate an additional pumparound. A part of the HVGO pumparound heat moves to a MVGO pumparound. This builds the HVGO pumparound temperature to a temperature higher than 600 ^{o}F. And reduces hot train preheat exchangers from six to three At the point when crude hydraulics are tight, a third pumparound can likewise enhance the reduction of the required HVGO pumparound flow rate to stay inside existing pump and piping limits. A third vacuum-unit pumparound frequently brings about the most minimal general cost arrangement

Conclusion

The ASPEN HYSYS was used to simulate the atmospheric distillation. this was done to improve cuts from petroleum and to improve the temperature of the residue channelled to the VDU. The results simulated indicated that volumetric flowrate of the atmospheric residue is sufficiently high while that of naphtha and other products had the low values of volumetric flowrate. This a clear indication that the outlet stream of the ADU has been improved. This improvement was achieved because the column was subjected to appropriated reflux efficiency in order to change more of the atmospheric residue to the premium cut like naphtha, kerosene, and diesel.

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