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Design and simulation of sour water unit: A case study of port Harcourt refining company

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ABSTRACT

The simulation of a 35.0m^3/hr sour water plant model using Aspen HYSYS process simulator was carried out using 3 different fluids; steam, kerosene and naphtha, respectively. The major equipment in the plant model are: 3-phase separator, heat exchanger and absorber tower. The results of the parameter obtained from aspen HYSYS were presented in terms of material balance, Energy balance, sizing and costing. The material balance result in terms of composition (mole fraction) obtained from the absorber outlet stream for treated water were constant for naphtha and kerosene plants showing 0.0248, 0.0004, 0.9748 and 0.0000 for NH_3, H_2S, H_2O and oil, respectively while that of steam plant gave 0.0296, 0.0004, 0.9699 and 0.0000 for NH_3, H_2S, H_2O and oil, respectively. The result in terms of material balance shows total elimination of oil from sour water with 97.48% recovery of water, 0.04% reduced rate of H_2S and 2.48% reduced rate of NH_3. The energy requirement of each plant model in terms of heat duty (kJ/hr) for heater and heat exchanger for steam, kerosene and naphtha plant models were also examined. Steam plant heat duty for heater and heat exchangers gave: 6.798 x 10^6 KJ/hr and 8.031 x 10^6 KJ/hr, respectively, kerosene plant heat duty for heater and heat exchanger were 8.031 x 10^6 KJ/hr and 2.067 x 10^7 KJ/hr, respectively, naphtha plant heat duty for heater and heat exchanger were 8.031 x 10^6 KJ/hr and 1.238 x 10^7 KJ/hr, respectively. Sizing results of major equipment such as 3-phase separator, heat exchanger and absorber column for steam, kerosene and naphtha plant were obtained from Aspen HYSYS simulation along with cost analysis.

Capsule Summary: Design and simulation of sour water was performed at port Harcourt refining company and study revealed that the naphtha plant found to be more economical and cheaper in terms of sour water design others.

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INTRODUCTION

In the petroleum refining process one of the main by-products is wastewater rich in Hydrogen sulphide and ammonia commonly referred to as sour water (Andy et al., 2011). The sour water generated in refineries is generally classified as being either phenolic or non-phenolic. Non-phenolic water contains almost exclusively NH_3, H_2S, and possibly a trace of CO_2; it is generated by refinery hydrotreating (hydrodesulphurization or HDS) units. When stripped of contaminants, non-phenolic water can typically be recycled for reuse in the HDS as wash water, or it can be used as makeup water to the crude desalting process (Weiland and Hatcher, 2012).

Sour water is produced in Atmospheric Crude Columns and Vacuum Crude Towers when stripping steam is condensed and removed by overhead condensing systems. It is also produced in Vacuum Crude Towers from equipment such as ejectors and barometric condensers which are designed to maintain vacuum inside the column. Steam injection to vacuum heater is another source of sour water in Vacuum Distillation Unit. In Thermal and Catalytic Cracking Units, sour water is produced as condensates from steam used in injection, stripping and aeration. Hydrotreater wash water is also another major source of sour water (Anonymous, 2023).

Heavy viscous feeds which are rich in sulphur produce high hydrogen sulphide concentrations when hydrogenated while ammonia is produced from hydrogenation of organic nitrogen compounds. If more sulphur will be removed to meet the more stringent environmental requirements, then there may be more nitrogen converted to ammonia which would accumulate in wash water. Hydrogen sulphide and ammonia concentrations are highest in sour water coming from Hydrodesulfurization Units and Fluid Catalytic Cracking Units. In addition to this, phenols are produced from reactions between steam and cyclic hydrocarbons. Sour water with extremely high concentration of phenols would come from the fluid catalytic units. Sour water systems are garbage disposals or toilets of refineries. It does not receive constant feed rate and composition. Any water-soluble waste produced in the refinery, either continuously, intermittently or in slug will be disposed into this system (Anonymous, 2023; Asquith and Moore, 2000).

Stripped sour water specifications for ammonia and hydrogen sulphide can be highly dependent upon the locale where the unit is installed and the final discharge requirements. Ammonia is harder to strip than hydrogen sulphide and typical targets for ammonia are 30-80 ppmw in the stripped water versus undetectable to less than 0.1 ppmw for hydrogen sulphide. Typical recent installations (Asquith and Moore, 2000; Quinlan and Hati, 2010) involve 35-45 actual trays with tray efficiencies quoted anywhere from 25 to 45%. In some cases, other alkaline contaminants besides ammonia may be present in sour water. Amine can carry over into the regenerator purge or it can be present from injection into the crude unit overhead for corrosion control. Sodium, potassium, and magnesium may also be present from impurities in the makeup water (hardness) or by water-contacting various products containing these compounds within the upstream units (Asai et al., 1990).

In upstream and downstream processes all kinds of waste oils/sludge are produced, the so-called “slop oils” or “oily sludge”. These products are generally difficult to manage, due to their high solids and emulsion content and high viscosity. The slop oils can have variations in their composition and can contain 10 – 70% oil, 30 – 90% water and 5 – 20% solids. The oil viscosity can be very high and reach values of up to 1000 cSt/ 50°C. The specific gravity of the oil (oil density) can vary from 0.8 to 0.98 g /ml (Aspen Technology, 2018).

Several contaminants such as H_2S, NH_3 and oil are present in sour water. The need to treat this water to eliminate the contaminant is very significant to every process industry because it is a potential hazard to the environment and its inhabitant. Base on the hazards, water from refinery operations cannot be discharge to sewer or reuse without treating the sour water. Therefore, this project seeks to design a sour water system that can reduce the concentration of H_2S, NH_3 and oil from refinery sour water. This shall be carried out using a simulation software involving three different stripping agents such as steam, naphtha and kerosene (Aspen Technology, 2018; Babich, 2003).

The aim of this study is to design and simulate a sour water plant that can reduce the concentration of NH_3, H_2S and oil using steam, naphtha and kerosene as stripping agents. The following objectives were carried in order to achieve this aim:

i. Study the process for the design of sour water using steam.
ii. Develop a process flow diagram for the design of sour water unit using naphtha.
iii. Develop a process flow diagram for the design of sour water unit using kerosene.
iv. Develop design equations for the dimensional and functional parameters of each of the unit operations and processes by the application of the principle of conservation of mass and energy.
v. Obtain process variables from relevant literature and incorporate it into the design performance equations.
vi. Use Aspen Plus to design, model and simulate the process plant
vii. Determine the dimensional and functional parameters using Aspen HYSYS and analytical method.
viii. Compare the effectiveness of steam, Kero and naphtha as stripping agents.
ix. Effectively cost the plant to know it fixed and operational cost.

MATERIAL AND METHODS

Sampling and methods

The materials used in this research work was the data from Port Harcourt Refining Company (PHRC) Sour water plant and includes the following: Detailed Process flow diagram, Inlet feed operating conditions, Comprehensive feed compositions, Utilities, Aspen HYSYS version 10 software, Chemical Engineering Handbooks. The design Assumptions/Constraints considered for the design of the sour water plants includes:

(i) The plant is operated at steady state.

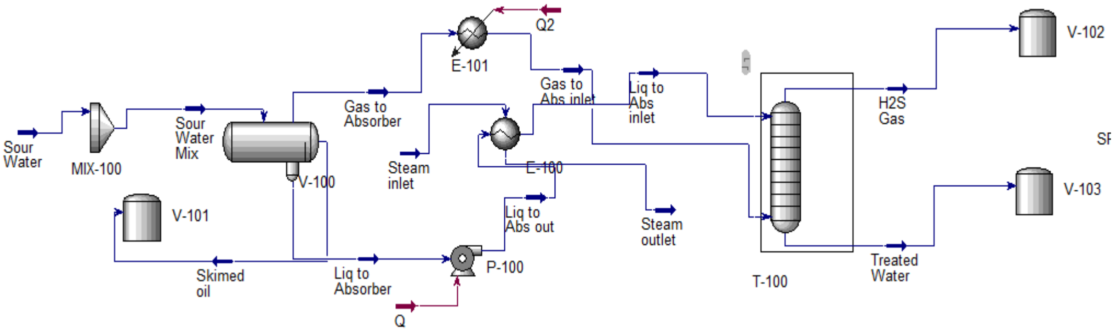

Fig. 1: Aspen HYSYS model of sour water plant using steam

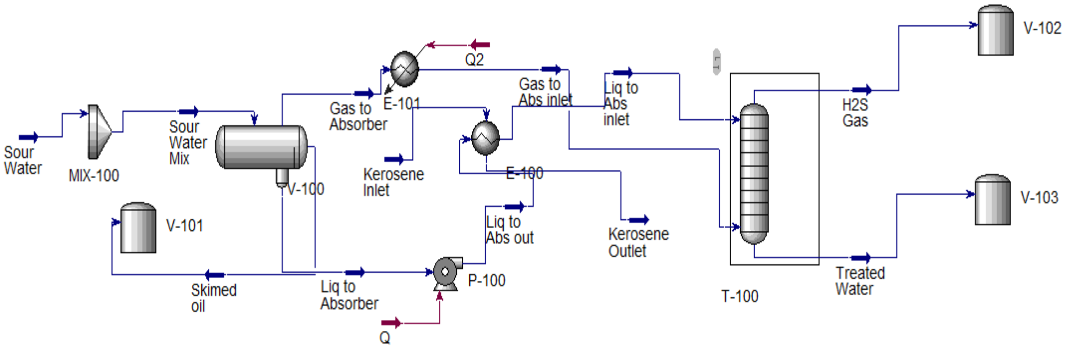

Fig. 2: HYSYS model of sour water plant using Kerosene

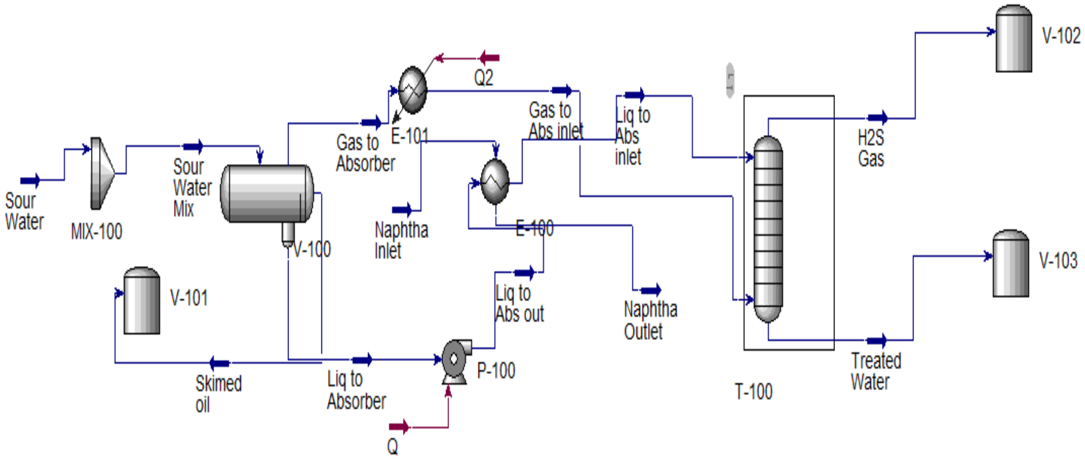

Fig. 3: HYSYS model of sour water plant using Naphtha

(ii) The composition of sour water changes along the trays of the absorber as absorption is occurring.
(iii) No reaction occurs as a result there is no generation or consumption of reactant species.
(iv) Balance is taken over entire volume of absorber
(v) The process is assumed to be isothermal.

The design of each unit in the plant would use the materials balance principle as stated in Eq. 1.

$$\begin{pmatrix}\text{Rate of accumulation}\\ \text{of material within the}\\ \text{unit operation}\end{pmatrix} = \begin{pmatrix}\text{Rate of inflow}\\ \text{of material into}\\ \text{the unit operation}\end{pmatrix} - \begin{pmatrix}\text{Rate of outflow}\\ \text{of material from}\\ \text{the unit operation}\end{pmatrix} \pm \begin{pmatrix}\text{Rate of generation/Consumption}\\ \text{of material due to chemical reaction}\end{pmatrix} \qquad (1)$$

Table 1: Result of the material balance for steam plant from HYSYS

Equipment	Parameters	Feed Composition		Output Composition		
3- Phase Separator	NH_3	0.0400		0.0684	0.0068	0.0368
	H_2S	0.0400		0.3888	0.0140	0.0005
	H_2O	0.8700		0.0511	0.0006	0.9627
	Slop oil	0.0500		0.4917	0.9785	0.0000
Heat Exchanger	NH_3	0.0368	0.0000	0.0368		0.0000
	H_2S	0.0005	0.0000	0.0005		0.0000
	H_2O	0.9627	1.0000	0.9627		1.0000
	Slop oil	0.0000	0.0000	0.0000		0.0000
Absorber	NH_3	0.0368	0.0684	0.1185		0.0296
	H_2S	0.0005	0.3888	0.3399		0.0004
	H_2O	0.9627	0.0511	0.1127		0.9699
	Slop oil	0.0000	0.4917	0.4290		0.0000

Table 2: Result of the material balance for kerosene and naphtha plant from HYSYS

Equipment	Parameters	Feed Composition		Output Composition		
3- Phase Separator	NH_3	0.0400		0.0684	0.0068	0.0368
	H_2S	0.0400		0.3888	0.0140	0.0005
	H_2O	0.8700		0.0511	0.0006	0.9627
	Slop oil	0.0500		0.4917	0.9785	0.0000
Heat Exchanger	NH_3	0.0368	0.0000	0.0368		0.0000
	H_2S	0.0005	0.0000	0.0005		0.0000
	H_2O	0.9627	1.0000	0..9627		1.0000
	Slop oil	0.0000	0.0000	0.0000		0.0000
Absorber	NH_3	0.0368	0.0684	0.1424		0.0248
	H_2S	0.0005	0.3888	0.3070		0.0004
	H_2O	0.9627	0.0511	0.1633		0.9748
	Slop oil	0.0000	0.4917	0.3872		0.0000

Table 3: Results of the energy balance for equipment

Plant	Equipment	Heat Duty (KJ/hr)
Steam	Heater	8031000
	Heat Exchanger	6798000
Kerosene	Heater	8031000
	Heat Exchanger	20670000
Naphtha	Heater	8031000
	Heat Exchanger	12380000

Table 4: Steam, naphtha and kerosene plant design specification for 3-phase separator 3-phase separator

Function	To separate the sour water into 3-Phaeses of light ends, medium liquids and heavy liquids			
Material Composition	Inlet	Output		
NH_3	0.0400	0.0684	0.0068	0.0368
H_2S	0.0400	0.3888	0.0140	0.0005
H_2O	0.8700	0.0511	0.0006	0.9627
Slop oil	0.0500	0.4917	0.9785	0.0000
Operating Condition				
Pressure	Min, 0KPag and Max, 103.425KPa			
Temperature	Min 34.34^{0}C and Max 121.1^{0}C			
Design Parameter				
Diameter	5.7912 metres			
Height	3.6576 metres			
Volume	96.3442m^3			
Power Source	Electricity			

Table 5: Kerosene plant design specification for absorber

Column tray	Trayed column			
Tray type	Bubble cap tray			
Function	To absorbs impurities such as ammonia, hydrogen sulphide gas in the sour water			
Material Composition	Inlet		Output	
NH_3	0.0368	0.0684	0.1424	0.0248
H_2S	0.0005	0.3888	0.3070	0.0004
H_2O	0.9627	0.0511	0.1633	0.9748
Slop oil	0.0000	0.4917	0.3872	0.0000
Operating Condition				
Pressure	Min, 0KPag and Max 103.425KPa			
Temperature	Min 46.830C and Max 121.110C			
Design Parameter				
Diameter	6.4008 metres			
Height	9.7536 metres			
Material	Stainless steel			
Power source	Electricity			

Table 6: Kerosene plant design specification for absorber

Column	Trayed column			
Tray type	Bubble cap tray			
Function	To absorbs impurities such as ammonia, hydrogen sulphide gas in the sour water			
Material Composition	Inlet		Output	
NH_3	0.0368	0.0684	0.1424	0.0248
H_2S	0.0005	0.3888	0.3070	0.0004
H_2O	0.9627	0.0511	0.1633	0.9748
Slop oil	0.0000	0.4917	0.3872	0.0000
Operating Condition				
Pressure	Min, 0KPag and Max 103.425KPa			
Temperature	Min, 43.750C and Max 121.110C			
Design Parameter				
Diameter	6.5532 metres			
Height	9.7536 metres			
Material	Stainless steel			
Power source	Electricity			

Table 7: Comparison of HYSYS simulation model results of steam plant with plant data for treated water

Components	Plant Data Composition (wt %)	HYSYS Model Composition (wt %)	Difference
NH_3	0.0045	0.0296	0.0251
H_2S	0.0004	0.0004	0.0000
H_2O	0.9951	0.9699	0.0252
Slop Oil	0.0000	0.0000	0.0000

Table 8: Comparison of HYSYS simulation model results of kerosene plant with plant data for treated water

Components	Plant Data Composition (wt %)	HYSYS Model Composition (wt %)	Difference
NH_3	0.0045	0.0248	0.0203
H_2S	0.0004	0.0004	0.0000
H_2O	0.9951	0.9748	0.0203
Slop Oil	0.0000	0.0000	0.0000

Design simulation process description

The simulation procedure started with opening a new case in Aspen HYSYS and selection of components for the simulation which were water, hydrogen sulphide and naphtha, the fluid package used for the model was Peng Robinson equation of state. The next step was to enter the simulation environment where the sour water plant model using steam, kerosene and naphtha, respectively was built, in the simulation environment the sour water stream was created and sent into a mixer before proceeding to a 3-phase separator, skimmed oil recovered from the 3-phase separator was sent into tank V-101 (slop tank) while the gas to absorber stream from the separator was heated with heater E-101 before entering the bottom part of the absorber. The last stream from the separator was pumped into heat exchanger E-100 where its temperature was raised to that of the gas stream before entering to the topmost part of the column (absorber) (Babich, 2003). The top product recovered from the absorber is H_2S gas because it contains more of hydrogen sulphide gas while the bottom product from the absorber is treated water because its impurities (H_2S, and NH_3 gas) has been removed. Both products recovered from the absorber tower were sent into their respective storage tanks. After the simulation was done using steam, it was also repeated using kerosene and Naphtha as shown in Figures 1 to 3. (Baukal et al., 2000).

Input simulation data

Sour water, normal design; volumetric flow rate (m^3/hr), 34.4, 35.0; expected temperature at sour water mix, 34.4 ^oC; feed pressure, 101.3kPa.

RESULTS AND DISCUSSION

Table 1 and 2 below shows the composition of material components balance for sour water plant using steam, naphtha and kerosene as stripping agents. The feed composition as obtained from port Harcourt reefing company limited enter the 3-phase separator through a mixer with 4% NH_3, 4% H_2S, 87% water and 5% slop oil for the individual plant. The separator remove hydrocarbon in the sour water mix through it water boot resulting in the recovery of 97.9% slop oil, 0.68% NH_3 while the gas to absorber stream from the separator having 38.9% H_2S and 49.17% slop passes through heater E-101 to the bottom of the absorber. The second outlet stream from the separator containing 96.27% water, 0.05% H_2S and 3.68% NH_3 passes through pump (P-100) and preheated in heat exchanger E-100 in order to raise it temperature to that of gas before entering the top of the absorber inlet. The output of the absorber contains a gas and treated water stream after absorption of the contaminant in the sour water (Bellen, 2009; Bravo et al., 2002; Beychock, 1967). The gas stream recovered 30.70% H_2S for kerosene and naphtha plant and 33.99% H_2S for steam plant. The treated water stream shows 97.48% recovery of water, 0.04% H2S, 2.48% NH_3 for kerosene and naphtha while that of steam plant gave 96.99% water, 2.96% NH_3, 0.004% H_2S and this in line with composition of treated water from Port Harcourt refining sour water plant (Butler, 1998; Ebrahimi et al., 2003; GEA, 2018; Port Harcourt Refinery, 1986).

Energy balance for sour water plant using steam, naphtha and kerosene as stripping agent

The general equation of conservation of energy can be written as shown in Eq. 2.

$$\begin{pmatrix}\text{Rate of inflow of energy}\\ \text{into the system}\end{pmatrix} = \begin{pmatrix}\text{Rate outflow of energy}\\ \text{from the system}\end{pmatrix} \pm \begin{pmatrix}\text{disappearance or generation}\\ \text{of Energy due to chemical}\\ \text{reaction}\end{pmatrix} + \begin{pmatrix}\text{Rate of accumulation}\\ \text{of energy within the}\\ \text{System}\end{pmatrix} \qquad (2)$$

Table 9: Comparison of HYSYS simulation model results of naphtha plant with plant data for treated water

Components	Plant Data Composition (wt %)	HYSYS Model Composition (wt %)	Difference
NH_3	0.0045	0.0248	0.0203
H_2S	0.0004	0.0004	0.0000
H_2O	0.9951	0.9748	0.0203
Slop Oil	0.0000	0.0000	0.0000

Table 10: Cost comparison of steam, kerosene and naphtha simulated plant models

		Cost (USD)	
Parameter	Steam Plant	Kerosene Plant	Naphtha Plant
Total Capital Cost	3534730	3556890	3533690
Total Operating Cost	1010870	1011940	1011760
Total Utilities Cost	65637.9	65637.9	65637.9
Equipment Cost	312000	319200	313800
Total Installed Cost	996600	1016300	998800

In the plant, the heating equipment used are fired heater furnace and heat exchanger, the fired heater furnace was modeled in Aspen HYSYS as a heater and its function is to pre-heat the sour water before it enters the absorber, the function of the heat exchanger is to raise the temperature of the gas that is entering the absorber. In the main absorber tower the difference in temperature of the incoming gas and liquid must be relatively close before the absorption efficiency can be maximum otherwise most of the gases will be lost through evaporation (Wang, 2012).

Table 3 below shows the energy requirement of the sour water plant using steam, Kerosene and naphtha as heating fluids. The heat flow as obtained from HYSYS shows that for steam plant, the heater (E-101) requires heat duty of 8031000KJ/hr, while the heat exchanger requires least heat flow of 6798000KJ/h. For kerosene plant, the heat duty required by the heater is 8031000KJ/hr and that of the heat exchanger is 20670000KJ/hr for it maximum operation while that of naphtha plant is 8031000KJ/hr for heater and 12380000KJ/hr for heat exchanger, respectively.

Equipment design specification for 3-phase separator for steam, kerosene and naphtha plants

Table 4 shows the design specification of 3-Phase separator used in the simulation of sour water plant. The separator separates the sour water mix into 3 – phases namely; light ends, medium liquids and heavy liquid. The separator operates at a temperature range of 34.34 ^{0}C to 121.1 ^{0}C and uses electricity as source of power. The table below summarizes the input, output, diameter, volume and height of the separator.

Equipment design specification of absorber for steam plant

Table 5 shows steam plant design specification of Absorber used in the absorption of H_2S, NH_3 and Oil from sour water. The equipment recovers 96.99% of H_2O with 0.04% of H_2S, 2.96% of NH_3 and 0% of Oil removed. This is shown in the table of steam plant Design specification for Absorber. Other Design parameters such as diameter, height, temperature and pressure are summarized in the table below.

Equipment design specification of absorber for kerosene plant

Table 6 shows kerosene plant design specification of Absorber used in the simulation of sour water. The equipment absorbed H_2S, NH_3, oil and recovered 97.48% of H_2O as output. The column is constructed with stainless steel and is the final stage in the simulation of sour water plant. It operates with temperature ranging from 46.83 ^{0}C to 121.11 ^{0}C and uses electricity as source of power. Other parameters including the input and outputs are summarized in Table 6 below.

Equipment design specification of absorber for kerosene plant

Table 7 shows the naphtha plant design specification for Absorber for the recovery of treated water from sour water before discharging to sewer or reuse in the refinery. The Absorber recovers 97.48% of H_2O and 0.04% of H_2S, 2.48% of NH_3 and 0% of oil. It operates with a minimum temperature of 43.75 ^{0}C and maximum of 121.11 ^{0}C with maximum pressure of 103.425 KPag. The Equipment is made of stainless steel and uses electricity as source of power. The input, output, diameter and other parameters as summarized below

Table 8 shows the difference between simulated result and plant data for steam plant. It can observe from the table that, there is a difference of 0.0251 mol% of Ammonia, 0.0000 mol% difference of hydrogen sulphide, 0.0252 mol% difference of water and 0.0000 mol% difference of oil from plant data. This shows that simulated results are not far from plant data.

Table 9 shows the difference between simulated result and plant data for kerosene plant. It can be observed from the table that, there is 0.0203 mol% difference of Ammonia, 0.0000 mol% difference of hydrogen sulphide, 0.0203 mol% difference of water and 0.0000 mol% difference of oil from plant data. This shows that simulated results are not far from plant data.

Table 10 shows the difference between simulated result and plant data for naphtha plant. It can be observed from the table that, there is 0.0203 mol% difference of ammonia, 0.0000 mol% difference of hydrogen sulphide, 0.0203 mol% difference of water and 0.0000 mol% difference of oil from plant data. This shows that simulated results are not far from plant data.

Cost comparison of steam, kerosene and naphtha simulated plant models

Table 11 shows the cost comparison of steam, naphtha and kerosene plant. Analyzing the table below, shows that steam plant is more economical having a total capital cost of $3,534,730: total operating cost of $1,010,870: equipment cost of $312,000 and total installed cost of $996,600 as against kerosene and naphtha plant.

Figure 4 shows how the composition (mole fraction) of water is changing along the trays of the absorber tower for steam, kerosene, and naphtha plant models, respectively. Observing Figure 4 carefully reveals that the composition of water is decreasing as the number of trays are increasing for steam, kerosene, and naphtha plant models, respectively, this can be attributed to the fact as more impurities (H_2S, and NH_3 gas) are been absorbed from the sour water so does its composition decreases which was initially very high with contaminants (H_2S, and

NH_3 gas), after the absorption of the contaminants from the sour water, the amount of concentration of contaminants left is now very small and of negligible amounts which can be discharged to the environment without causing any contamination to the environment.

Figure 5 shows how the composition (mole fraction) of ammonia is changing along the trays of the absorber tower for steam, kerosene, and naphtha plant models, respectively. Observing Figure 5 carefully reveals that the composition of ammonia is increasing as the number of

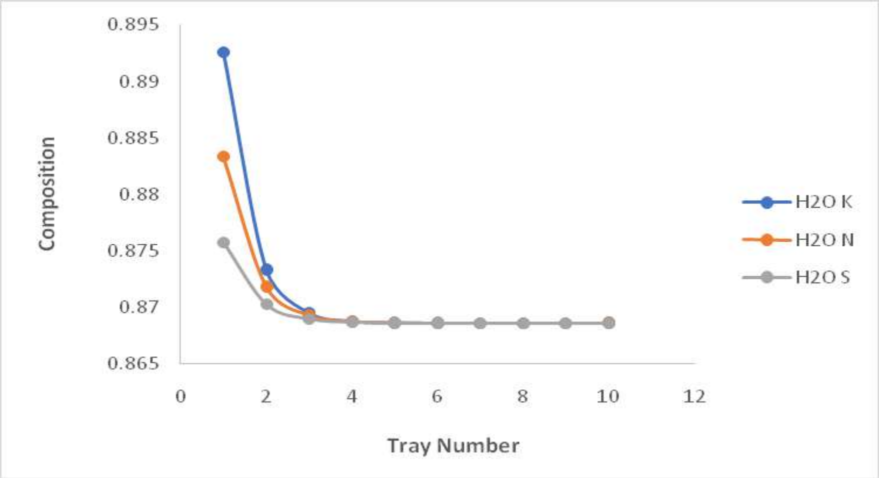

Fig. 4: Plot of H_2O composition for kerosene, naphtha and steam plant models changing with number of trays

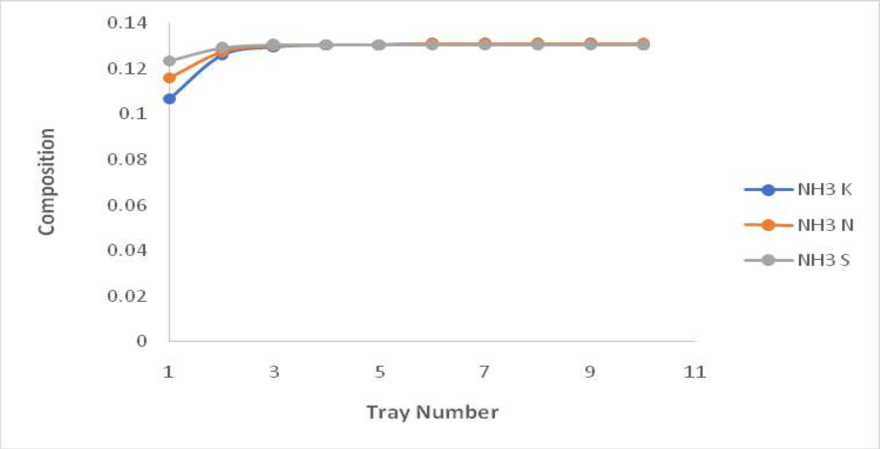

Fig. 5: Plot of NH_3 composition for kerosene, naphtha and steam plant models changing with number of trays

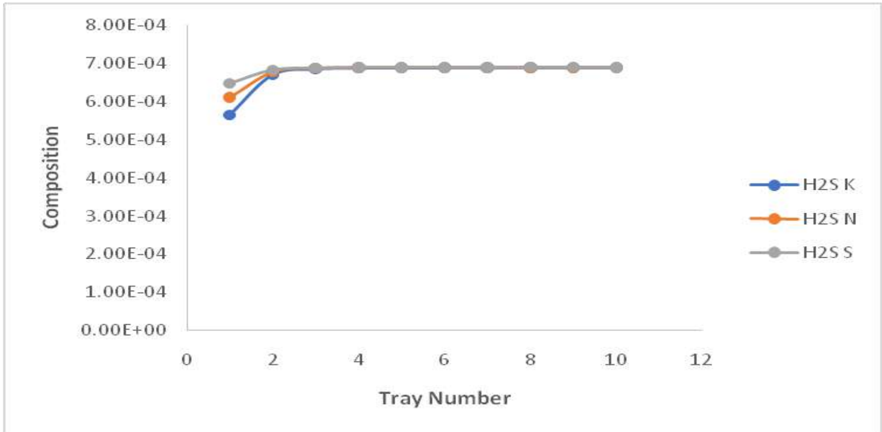

Fig. 6: Plot of H_2S composition for kerosene, naphtha and steam plant models changing with number of trays

trays are increasing for steam, kerosene, and naphtha plant models, respectively, this can be attributed to the fact as more impurities (H_2S, and NH_3 gas) are been absorbed from the sour water so does its composition decreases which was initially very high with contaminants (H_2S, and NH_3 gas), after the absorption of the contaminants from the sour water, the amount of concentration of contaminants left is now very small and of negligible amounts which can be discharged to the environment without causing any contamination to the environment.

Figure 6 shows the concentration of hydrogen sulphide is highest for steam plant model since it is the base case simulation followed by naphtha and kerosene plant models. The concentration of all three plant models decreases exponentially initially but later appears to be steady afterwards because the maximum absorptivity has been exceeded hence it can no longer absorb any more contaminants at this value so it continues linearly throughout the trays.

CONCLUSIONS

The simulation of Sour water plant using steam, kerosene and naphtha in Aspen HYSYS software version 10, was carried out. The parameters obtained are; balance, sizing, costing and sensitivity analysis. The major equipment used in the simulation were: 3-phase separator for separating the sour water feed into light, medium and heavy components, heat exchanger for raising the temperature of the liquid before entering the absorber and absorber for absorbing impurities H_2S, NH_3 and oil from the sour water. The material balance for naphtha and kerosene plant shows a constant composition (mole fraction) of the absorber outlet for treated water stream as 0.0248, 0.0004, 0.9748 and 0.0000 for NH_3, H_2S, H_2O and oil, respectively. That of steam plant gave 0.0296, 0.0004, 0.9699 and 0.0000 for NH_3, H_2S, H_2O and oil, respectively. The result obtained show a complete elimination of oil from sour water, with 97.48% recovery of water, 2.48% reduction rate of NH_3 and 0.04% reduction rate of H_2S for both naphtha and kerosene plants. The energy requirement of each plant model in terms of heat duty (KJ/hr) for heater and Heat exchanger for steam, kerosene and naphtha plant models were also examined as steam plant heat duty for heater and heat exchangers gave: 8.031×10^6 KJ/hr and 6.798×10^6 KJ/hr, respectively. Kerosene plant heat duty for heater and heat exchanger were 8.031×10^6 KJ/hr and 2.067×10^7 KJ/hr, respectively. Naphtha plant heat duty for heater and heat exchanger were 8.031×10^6 KJ/hr and 1.238×10^7 KJ/hr, respectively. The cost for each of the plant also show a total capital cost of $3,534,730 for steam plant, $3,533,690 for naphtha plant and $3,556,890 for kerosene plant. This implies naphtha plant is more economical and cheaper in terms of sour water design.

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