
DESIGN AND FABRICATION OF A 5 BSPD PILOT-SCALE SEMI-REGENERATIVE CATALYTIC NAPHTHA REFORMING UNIT

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ABSTRACT

Catalytic naphtha reforming process is a vital refinery process where low octane naphtha is converted into high octane reformate while concurrently generating hydrogen gas. This research work presents the design, simulation, and fabrication of 5 barrels per stream day (BPSD) pilot-scale semi-regenerative catalytic naphtha reforming process which serves as part of the mini-refinery development process at the Ahmadu Bello University (ABU), Zaria. The design was based on Escravos Heavy Hydrotreated Naphtha serving as feed; with a capacity of 5 barrel/day. Material and Energy balances calculations were performed on the developed process flow diagram before detailed equipment design was done using Mathcad and Aspen Hysys simulation. The designed equipment that constitute the pilot plant are: Surge Drum (D-01), Heat Exchanger (EX-01), Phase Separator (D-02), three Fired Heaters (H-01, H-02 & H-03) and three Fixed Bed Reactors (R-01, R-02 & R-03). R-01 (0.02 m³ volume and 805.2 mm height), R-02 (0.03 m³ volume and 923.0 mm height) and R-03 (0.04 m³ volume and 1203.0 mm height).

Keywords: Catalytic reforming; pilot-scale refinery; heavy naphtha; process design; reformate

1. INTRODUCTION

Naphtha reforming by means of catalysts is an extremely significant conversion reaction in the current petroleum refining plants since it involves upgrading of the heavy naphtha with low octane ratings to high-octane gasoline fractions while generating the hydrogen-containing gas at the same time (Gary et al., 2007; Speight, 2020; Towler & Sinnott, 2022). Apart from that, besides being beneficial for the production of gasoline, the catalytic naphtha reforming is one of the main sources of BTX aromatic compounds, which are vital for petrochemical industry. Hydrogen produced during the catalytic reforming of the naphtha is also essential since it acts as a reactant in the hydrocracking and hydrotreating reactions (Aznárez et al., 2024; Saeidi et al., 2014).

The main feedstocks used in catalytic reforming are heavy naphthas composed mainly of paraffins, naphthenes, and aromatics having hydrocarbon molecular weight from C₆ to C₁₂ which usually boil within the temperature range of 90–200 °C (Speight, 2014). The feedstocks are pre-treated before entering the reforming unit using hydrotreatment process, where sulfur, nitrogen, and oxygen containing compounds which might inhibit platinum reforming catalysts activity are removed (Jones & Pujadó, 2006; Speight, 2020). Bifunctional catalysts containing platinum supported on chlorinated alumina act as a catalytic material in reforming process; metallic centers in the catalyst provide hydrogenation/dehydrogenation reactions whereas acid sites enable isomerization and cyclization reactions.

Major reactions taking place during the reforming process involve naphthenes dehydrogenation to aromatics, paraffins dehydrocyclization, paraffins and cycloparaffins isomerization, and minimal hydrocracking. All the above reactions result in increase in the rating of gasoline octane number through production of hydrogen gas which is an important industrial gas. Catalytic reforming process is usually categorized into semi-regenerative reforming (SRR), cyclic catalytic reforming (CCRY) and continuous catalyst regeneration (CCR) processes based on the catalyst regeneration technique used in the process

(Bell, 2001). Out of all the mentioned industrial catalytic reforming processes, the semi-regenerative reforming is one that has remained attractive for small and medium capacity refineries due to its simple process design, low capital cost and easy operation when compared to continuous catalyst regeneration processes (Bell, 2001; Gary et al., 2007). Even though continuous catalyst regeneration offers increased yield of reformate and increased catalyst life, semi-regenerative reforming process is still widely used in pilot plants, modular refineries and educational institutions because of its economic viability (Towler & Sinnott, 2022).

Modern trends in catalytic reforming have mainly focused on advanced platinum based catalysts, catalyst regeneration, molecular kinetic modelling, reactor optimisation, increasing hydrogen yield and process intensification (Ruban et al., 2026; Velázquez-Alonso et al., 2025; Aznárez et al., 2024; Ali et al., 2023; Zhang et al., 2022). While these developments are generally important, comparatively less attention has been given to the practical design and fabrication of catalytic reforming unit. This aspect is especially important in developing countries such as Nigeria, where limited local refinery equipment manufacturing capacity, high dependence on imported process equipment and the cost of acquiring sophisticated commercial-scale units can constrain research, training, technology development and local process innovation. On the other hand, process simulation software such as Aspen HYSYS has become an essential engineering tool for refining processes design and optimization, by allowing rigorous calculations for material and energy balances, equipment sizing, heat integration, and evaluation of operating conditions prior to plant construction (Ritov & Dana, 2026; Pérez et al., 2025; Towler & Sinnott, 2022). These developments have greatly reduced design uncertainties and improved the efficiency and reliability of refinery process design.

In spite of these developments, most of the published studies have focused on catalyst synthesis, process optimization and mathematical modelling whereas relatively less attention has been paid to the indigenous design and fabrication of pilot scale catalytic reforming units, particularly in the developing countries. The existence of pilot scale reforming facilities is crucial for validation of engineering designs, catalyst performance evaluation, operational studies, student training and bridging the chasm between laboratory scale investigations to commercial refinery operations (King, 2013; Meyers, 2004). Indigenous pilot plants are also useful platforms for technology demonstration, process development and capacity building in petroleum refining.

The Ahmadu Bello University (ABU) Mini-Refinery Development Project was conceived to strengthen petroleum engineering education and research through the establishment of integrated pilot-scale refinery process units. As part of this initiative, a catalytic reforming unit is required to upgrade hydrotreated heavy naphtha into high-octane reformate suitable for gasoline blending while simultaneously producing hydrogen for downstream refinery operations. Therefore, this study presents the design, simulation, and fabrication of a 5 barrel per stream day (BPSD) pilot-scale semi-regenerative catalytic naphtha reforming unit using Escravos heavy hydrotreated naphtha as the design feedstock. The work covers process flow development, material and energy balance calculations, equipment sizing using Aspen HYSYS and Mathcad, mechanical design, fabrication, and assembly of the pilot-scale reforming unit. The developed facility provides an indigenous platform for engineering education, catalyst evaluation, process optimization, and future refinery research in Nigeria.

2. METHODOLOGY

This chapter presents details of activities performed for the design and fabrication of the Pilot Scale Naphtha Catalytic Reforming Unit.

2.1 Project Execution Stages

The design was executed in stages as presented in Figure 1. The methodology outlined twelve (12) stages for the design and fabrication of the pilot plant.



Figure 1: Project Flow Diagram

2.2 DESIGN BASIS AND FEEDSTOCK ANALYSIS

The feed (Heavy Naphtha) was first analysed before establishing a design basis. The design basis and feedstock composition for the CRU are shown in Tables 1 and 2 respectively.

Table 1: CRU Design Basis

NHU Capacity	5 barrel/Day ($3.31 \times 10^{-2} \text{ m}^3/\text{h}$)
Feed Type	Escravos SR Heavy Naphtha
Source	Naphtha Hydrotreating Unit
Specific Gravity	0.78

Table 2: CRU Feedstock Composition (Escravos Assay, 2012)

S/No.	Components	Mole Fraction	Percentage(%)	Molecular Weight
1.	Paraffins	0.364	36.4	72
2.	Naphthenes	0.392	39.2	92
3.	Aromatics	0.244	24.4	84
Total/Average		1.000	100	82.67

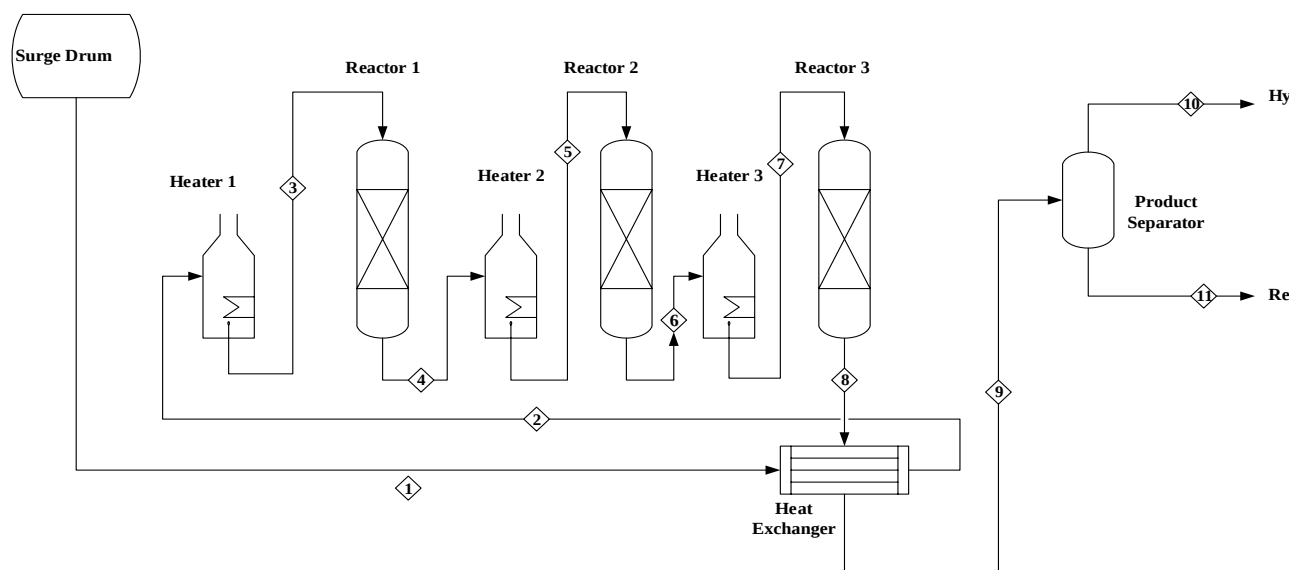


Figure 2: Process Flow Diagram for Semi Regenerative Catalytic Naphtha Reformer

2.3 PROCESS DESCRIPTION

The flow diagram of the catalytic naphtha reforming unit modeled in this work is displayed in Figure 2. The feed stock to the reformer is the heavy straight-run hydrotreated naphtha with boiling points between 90 °C and 200 °C and consists of molecules with 6–12 carbons. The catalyst mostly used in reforming operation is a bifunctional metallic catalyst typically - Pt-Re/Al₂O₃, providing the metal site and the acid site. In the initial stage, the desulfurized feed is mixed with hydrogen (in a mixer before preheating in heat exchanger (EX-01). The preheated feed, after mixing with hydrogen, is passed through fired heater H-01, where its temperature is raised to the required reaction temperature of 450–550 °C before entering the first catalytic reactor, R-01. The effluent from R-01 is then directed to fired heater H-02, where it is reheated to the required reaction temperature before entering the second reactor, R-02. Similarly, the effluent from R-02 is passed through fired heater H-03 to restore the temperature required for reaction before entering the third reactor, R-03. The use of interstage heaters is necessary because the principal reforming reactions are predominantly endothermic and consequently result in a reduction in temperature as the reaction proceeds through each reactor. The effluent from R-03 subsequently passes through the feed/product heat exchanger, EX-01, where heat is recovered to preheat the incoming fresh feed. The cooled effluent then enters the phase separator, D-02, where the hydrogen-rich gas phase is separated from the liquid reformate. The hydrogen-rich gas may then be recycled and mixed with the fresh feed upstream of R-01, while the liquid reformate is withdrawn as the bottom product.

2.4 EQUIPMENT DESIGN

2.4.1 Design Equations

The following were the equations used in the design of the various components of the CRU.

1. REACTORS

- REACTOR DESIGN EQUATION (PACKED BED REACTOR)
- REACTOR PRESSURE DROP EQUATIONS

$$\Delta P_B = 150 \cdot \frac{(\mu \cdot v_s)}{(D_p)^2} \cdot \frac{(1 - \epsilon)^2}{(\epsilon)^3} + 1.75 \cdot \frac{[\rho \cdot (v_s)^2]}{[D_p]} \cdot \frac{(1 - \epsilon)}{(\epsilon)^3}$$

Where

μ = average viscosity at inlet condition

v_s = superficial viscosity

D_p = catalyst particle size

ρ = catalyst particle density

ϵ = bed void space

$$W = \int_{F_{Ao}}^{F_{A1}} \frac{1}{r_A} dF_A$$

(1)

- TRANSPORT EQUATIONS

$$LHSV = \frac{\text{Mass_flowrate_inlet}}{\text{Mass_catalyst}}$$

(2)

- REACTOR PRESSURE DROP EQUATIONS

$$\Delta P_B = 150 \cdot \frac{(\mu \cdot v_s)}{(D_p)^2} \cdot \frac{(1 - \epsilon)^2}{(\epsilon)^3} + 1.75 \cdot \frac{[\rho \cdot (v_s)^2]}{[D_p]} \cdot \frac{(1 - \epsilon)}{(\epsilon)^3}$$

Where

μ = average viscosity at inlet condition

v_s = superficial viscosity

D_p = catalyst particle size

ρ = catalyst particle density

ϵ = bed void space

(3)

2. HEAT EXCHANGER

(4a)

Where

Q = Heat Transfer Rate
 U = Overall Heat Transfer Coefficient
 A = Heat Transfer Area
 Th_1 = Hot fluid Inlet
 Th_2 = Hot fluid outlet
 Tc_1 = Cold fluid Inlet
 Tc_2 = Cold fluid outlet
 D = Heat transfer diameter (mm)
 L = Heat transfer Length (mm)
 Re = Reynolds No.
 μ = Viscosity
 ρ = Density (kg/)
 V = Velocity (m/s)
 k = Thermal Conductivity
 Cp = Specific heat capacity

$$Q = UA\Delta T_m$$

$$A = \frac{\pi D^2 L}{4}$$

$$\Delta T_m = \frac{(Th_2 - Th_1) - (Tc_2 - Tc_1)}{\ln \left[\frac{Th_2 - Th_1}{Tc_2 - Tc_1} \right]}$$

$$Re = \frac{\rho V D}{\mu}$$

$$Pr = \frac{Cp \mu}{K}$$

(4b)

(4c)

(4d)

(4e)

3. FIRED HEATER

$$q_r = 0.173 f \left[\left(\frac{T_G}{100} \right)^4 - \left(\frac{T_S}{100} \right)^4 \right] \alpha_{CP} A_{CP} + hA(T_G - T_S)$$

$$N_{tubes} = \frac{Q}{2\pi r L q} \text{ where } r = d_o / 2$$

$$A_{CP} = ctc \times L \times N$$

Where

f - overall exchange factor.

T_G - temperatures of the flue gases leaving the radiant section

T_S - surface temperature of the tubes.

A_{CP} - equivalent cold plane surface area, ft^2 ,

α_{CP} is a factor by which A_{CP} must be reduced to obtain effective cold surface.

$h \approx 2$, $A \approx 2\alpha_{CP} A_{CP}$ and $f \approx 0.57$

(5a)

(5b)

(5c)

2.4.2 Summary of Material and Energy Balance

The material balance was established for the steady-state design basis of the 5 BPSD catalytic reforming unit. The calculated feed rate to the reforming section was 25.72 kg/h, comprising 8.57 kg/h of paraffins, 10.54 kg/h of naphthenes, 6.51 kg/h of aromatics and 0.10 kg/h of hydrogen. The component balances across the three (3) reactors show the progressive conversion of paraffins and naphthenes to aromatics, accompanied by hydrogen generation. The principal material-balance results for the reactor train and final product separation are summarized in Table 3.

Table 3: Summary of material balance across the catalytic reforming unit

Stream/Unit	Paraffins (kg/h)	Naphthenes (kg/h)	Aromatics (kg/h)	H ₂ (kg/h)	Total (kg/h)
Feed to CRU	8.57	10.54	6.51	0.10	25.72
R-01 outlet	3.43	5.40	11.65	5.24	25.72
R-02 outlet	1.37	3.34	13.71	7.30	25.72
R-03 outlet	0.55	2.52	14.53	8.12	25.72
Separator gas product	—	—	—	7.714	8.0609
Separator liquid reformat	0.5445	2.4696	14.240	0.7529	17.6601

However, the computed mass flow of the total continues to be nearly 25.72 kg/h in the reactor train. On the other hand, the mass flow of the components varies increasingly due to reforming reactions. For instance, aromatics rise from 6.51 kg/h in the feed stream to 14.53 kg/h on the effluent side of R-03. Similarly, paraffins and naphthenes decrease from 8.57 and 10.54 kg/h to 0.55 and 2.52 kg/h respectively. Hydrogen rises from 0.10 kg/h in the feed to 8.12 kg/h on the effluent side of R-03, which is indicative of hydrogen formation as a result of dehydrogenation reactions.

Table 4: Summary of calculated energy requirements for the major process units

Equipment	Function	Calculated energy requirement / enthalpy change (MJ/h)
EX-01	Feed/product pre-heater	74.62
H-01	Feed heating before R-01	201.58
R-01	Endothermic catalytic reforming	8,289.82
H-02	Interstage heating before R-02	1,088.55
R-02	Endothermic catalytic reforming	11,787.37
H-03	Interstage heating before R-03	11,800.65
R-03	Endothermic catalytic reforming	11,797.63

The energy requirements presented in Table 4 were determined from the enthalpy differences between the inlet and outlet process streams for each unit. The calculated heat-transfer duties for EX-

01, H-01, H-02 and H-03 were 74.62, 201.58, 1,088.55 and 11,800.65 MJ/h, respectively. For the reactors, the calculated enthalpy changes associated with the endothermic reforming reactions were 8,289.82, 11,787.37 and 11,797.63 MJ/h for R-01, R-02 and R-03, respectively. These values provided the thermal basis for the design of the heat-transfer equipment and the assessment of the energy requirements of the three-stage reforming process.

2.4.3 Numerical Design Outputs

The major equipment involved in the design comprises the surge drum, feed-to-product heat exchanger, three fixed-bed catalytic reactors, three interstage fired heaters and a phase separator. The calculated reactor volumes were 0.02, 0.03 and 0.04 m³ for R-01, R-02 and R-03, respectively, with corresponding catalyst bed lengths of 277, 417 and 695 mm, and overall heights of 805.2, 923.0 and 1203.0 mm respectively. The calculated fired heater duties for H-01, H-02 and H-03 were 201.58, 1088.55 and 11,800 MJ/h, respectively. The phase separator was designed as a vertical cylindrical vessel with a diameter of 203.2 mm, an overall height of 900 mm and a gross internal volume of 0.0281 m³. The summary of the calculated design parameters is presented in Table 5 and provides the basis for preparation of the equipment drawings and fabrication of the pilot-scale catalytic reforming unit.

Table 5: Summary of calculated design parameters of the 5 BSPD catalytic reforming unit

Equipment	Design parameter	Value
Surge drum D-01	Volume	0.20 m ³
	Diameter	500 mm
	Length	1000 mm
Heat exchanger EX-01	Type	Double-pipe, counter-current
	Tube diameter	19.4 mm
	Shell diameter	205 mm
	Heat-transfer area	0.036 m ²
Reactor R-01	Volume	0.02 m ³
	Overall height	805.2 mm
	Catalyst-bed length	277 mm
Reactor R-02	Volume	0.03 m ³
	Overall height	923.0 mm
	Catalyst-bed length	417 mm
Reactor R-03	Volume	0.04 m ³
	Overall height	1203.0 mm
	Catalyst-bed length	695 mm
Fired heater H-01	Heat duty	201.58 MJ/h
Fired heater H-02	Heat duty	1088.55 MJ/h
Fired heater H-03	Heat duty	11,800 MJ/h
Phase separator D-02	Diameter	203.2 mm
	Overall height	900 mm
	Gross volume	0.0281 m ³

2.5 WORKING DRAWINGS / MODELS

Figures 3 – 9 presents the working drawings and models based on the equipment design calculations. The drawings were very essential as they served as working templates for fabrication.

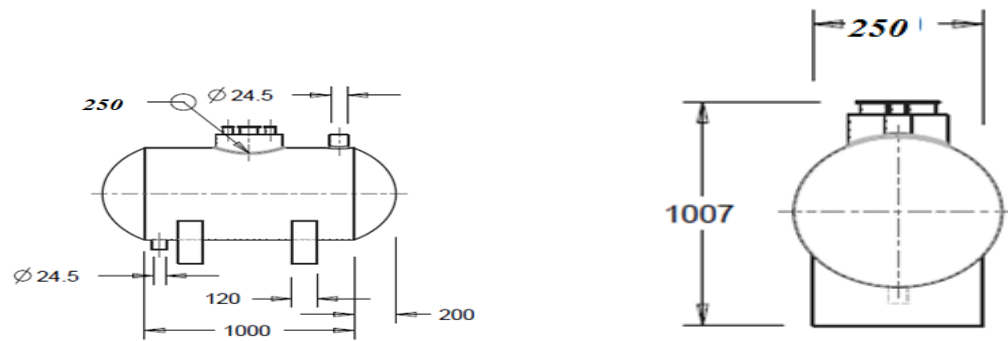


figure 3: Working Drawing of Surge Drum

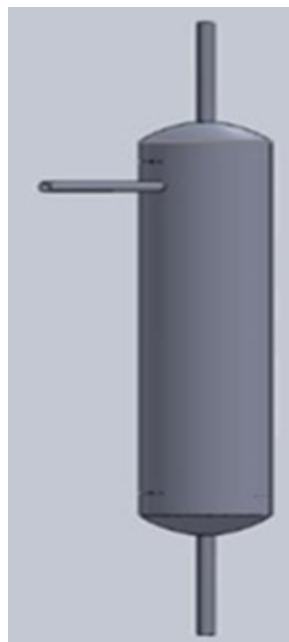
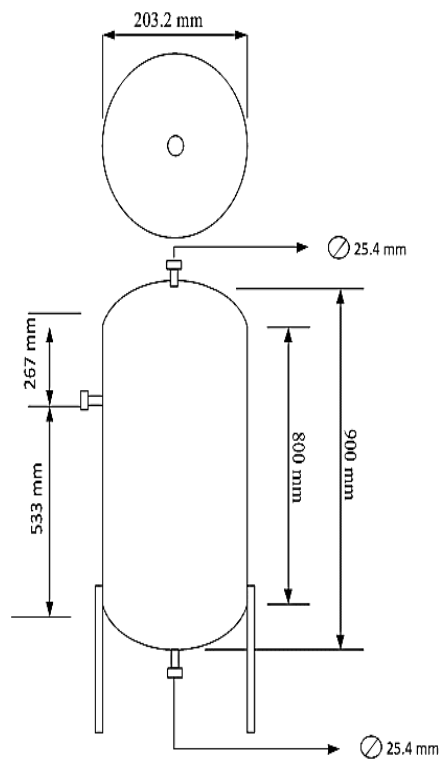


Figure 4: Working Drawing/Models of Phase Separator

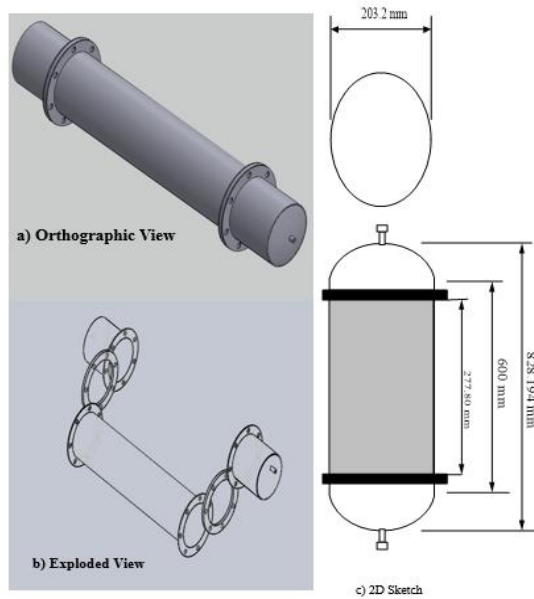


Figure 5: Working Drawing/Models of CRU Reactor (01)

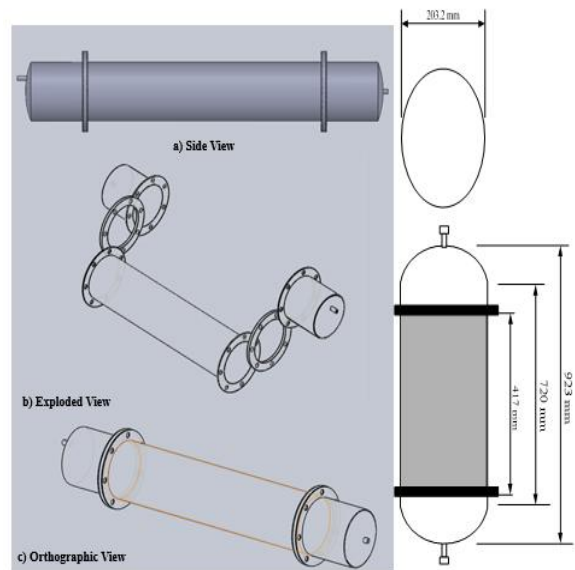


Figure 6: Working Drawing/Models of CRU Reactor (02)

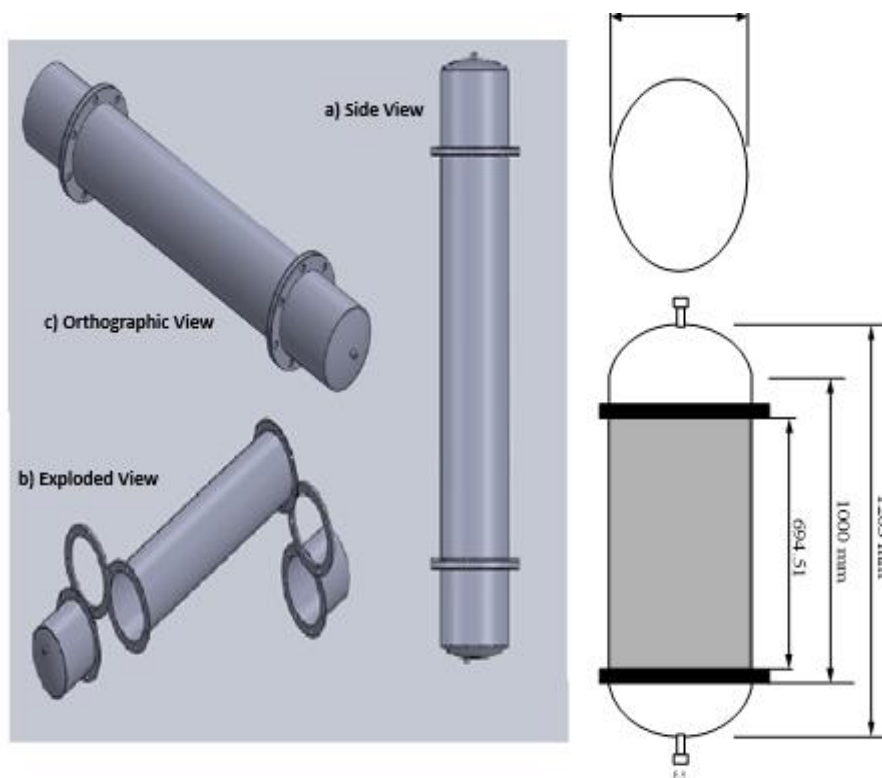


Figure 7: Working Drawing/Models of CRU Reactor (03)

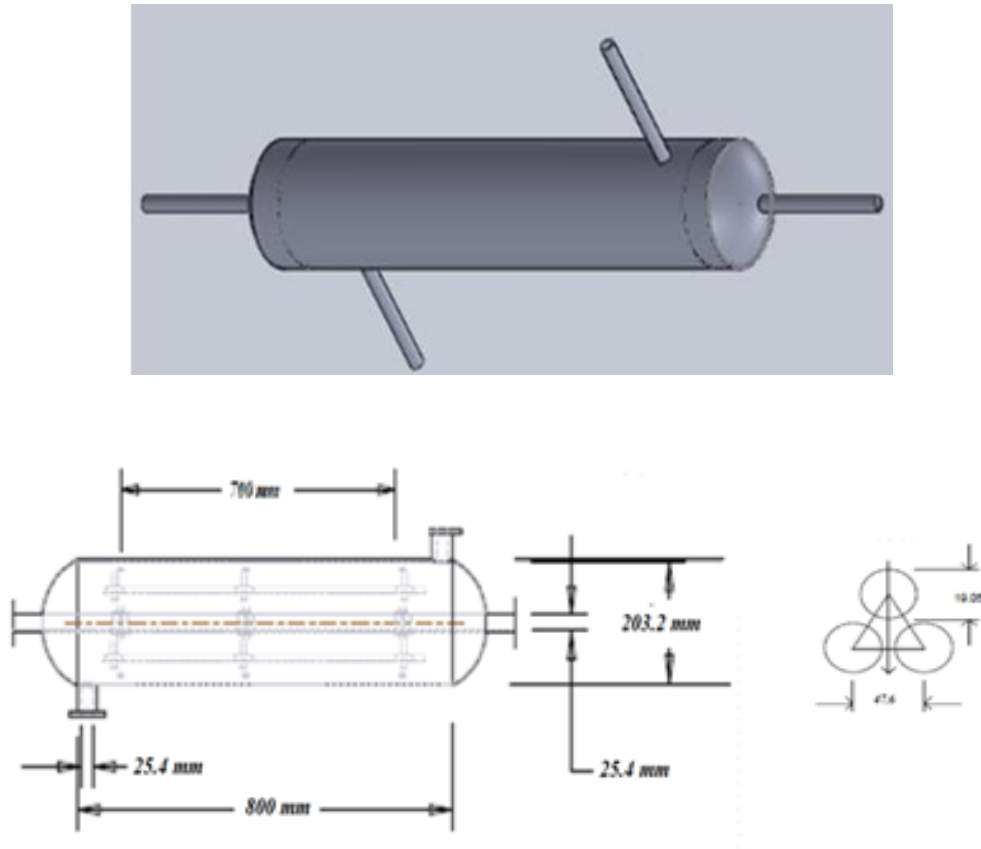


Figure 8: Working Drawing/model of Heat Exchanger

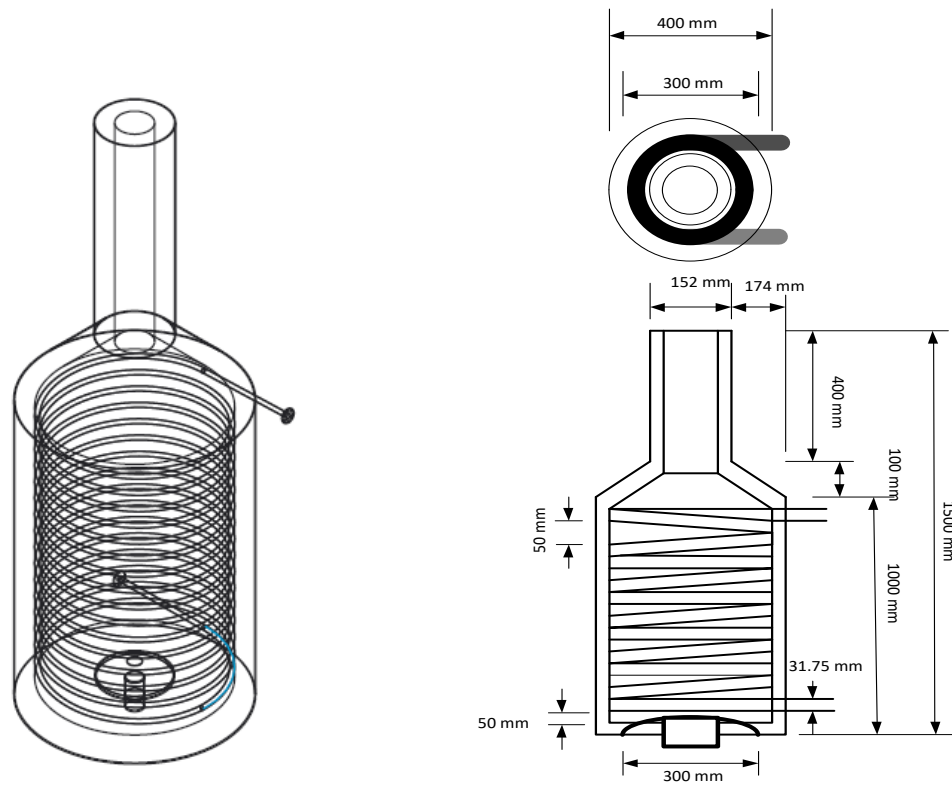


Figure 9: Working Drawing of Fired Heater

2.6 ECONOMIC CONSIDERATIONS, LOCAL FABRICATION AND SCALE-UP POTENTIAL

The fabrication of the 5 BPSD catalytic reforming unit was performed keeping in mind the use of locally available engineering materials, shop facility and fabricating skills. Some of the major equipment fabricated were the surge drum, feed/product heat exchanger, three fixed bed catalytic reactors, three fired heaters and the phase separator. This helped reduce the need for imported proprietary pilot scale equipment and proved the practicability of transforming the process design to a functioning pilot plant by utilizing local engineering and fabrication skills.

Economically speaking, there are advantages associated with local fabrication of equipment for pilot-scale petroleum processing studies because it helps in designing and manufacturing the equipment according to the needs of the specific application and not purchasing as imported packages. It also helps in modifying, maintaining and replacing parts of the apparatus as well as upgrading the apparatus in accordance with further experimental requirements. But, the current study focused mainly on the technical design and fabrication of the unit and a detailed techno-economic study or a life cycle cost analysis of the items used was out of its scope. Hence, no general statement is made about the cost advantage of the fabricated unit over the imported commercial equipment.

The local fabrication technique also offers a way for future scaling up and commercialization of the system. The current 5 BPSD unit serves as a platform for generating operating data, performance of equipment and fabrication experience before proceeding onto building large units. Future scaling up must maintain the same process arrangement developed in the current study and include proper scaling changes in the size of equipment, heat transfer capabilities, sizing of reactors, instrumentation and control of processes as well as material used. It must also be accompanied by a detailed capital cost estimation, cost comparison of fabrication, energy consumption and economic feasibility study. Such studies will establish the commercial feasibility of locally made catalytic reforming systems and the identification of components that can be locally made and those that require special procurement.

3. RESULTS AND DISCUSSION

3.1 Equipment Fabrication

All the three (3) reactors are fabricated using stainless steel though mild steel was used in other units such heat exchanger and in phase separator as opposed to stainless steel specified by Aspen Hysys as the material of construction. The equipment was originally specified for fabrication from stainless steel; however, mild steel was used for selected components during fabrication because of cost and local material availability. The use of mild steel for appropriately selected refinery and process equipment is consistent with established industrial practice world wide, Accordingly, the use of mild steel in selected components of the pilot unit does not in itself constitute a process-design deficiency, although the material must remain compatible with the operating environment and be appropriately protected and monitored. The fabricated units are presented in Plate 1-4.



Plate 1a : Parts of the three Fabricated Reactors



Plate 1b : The three Fabricated Reactors

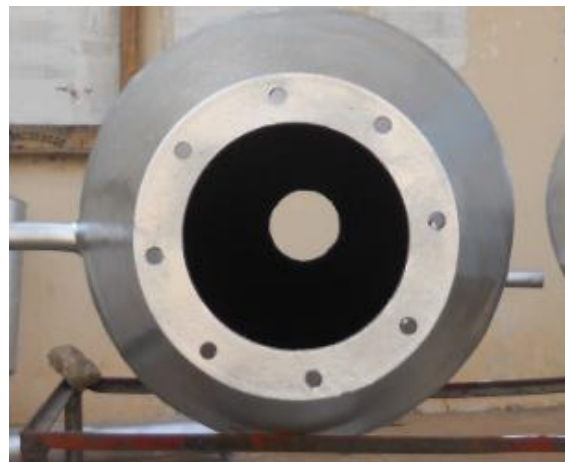


Plate 2: CRU Fired-Heater



Plate 3a: Heat Exchanger bundles



Plate 3b: Heat Exchanger tube



Plate 4: Surge Drum

3.2 ASSEMBLY/ INSTALLATION OF CRU PLANT

This section presents the installation process for putting up the unit. This involves the installation of all fabricated units, piping, as well as installation of controls such as flow valves, pressure gauges and thermocouples. It also covers the installation of three burners for the fabricated fired heaters. All these are presented in Plate 5 and the model of the installed unit is depicted using SOLID WORKS 2016 in Figure 10 (a & b).

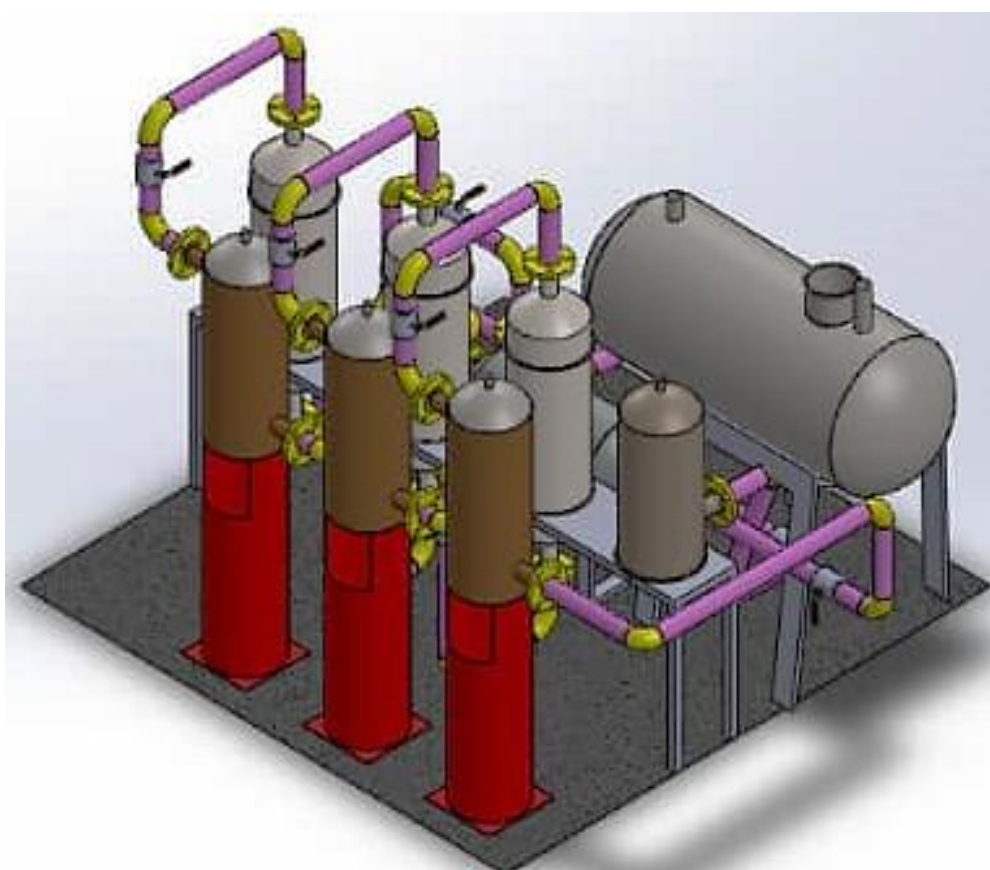
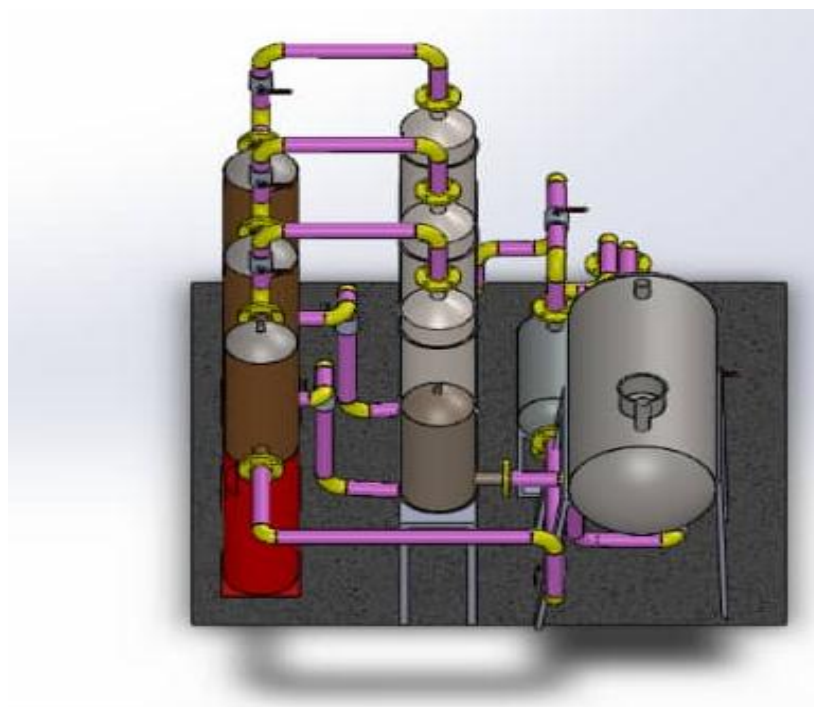


Figure 10 (a & b): 3D Diagram of the installed CRU



Plate 5: Cladding and Lagging

3.3 RON and MON of Fabricated CRU Processed Naphtha

The final experimental operation was conducted at a reactor temperature of between 460 – 480 °C and an operating pressure of approximately 3 bar and a feed rate was approximately 22.2 kg/h (0.5 L/min). Although hydrogen addition was included in the design and simulation basis, no external hydrogen was introduced during the actual test run. These conditions differed from the original design simulation basis.

To actually have a basis for interpreting the experimental results, a comparison with the design parameters is presented in Table 6.

Table 6: Comparison of design/simulation and actual test-run operating conditions

Parameter	Design/Simulation	Actual Test Run
Reactor temperature	450 – 550 °C	460 – 480 °C
Operating pressure	6 bar	Approximately 3 bar
Feed rate	25.72 kg/h (0.58 L/min)	22.2 kg/h (0.50 L/min)
Hydrogen addition	0.10 kg/h	None

The actual operating conditions represent the conditions recorded during the experimental test run, whereas the design/simulation values represent the original process-design basis. The comparison presented in Table 6 shows that, although the experimental reactor temperature remained within the design temperature range, the unit was operated at a substantially lower pressure and feed rate than the design basis, and no external hydrogen was introduced. These differences are important when interpreting the experimentally obtained reformate quality because the test run did not reproduce all of the conditions used for the original design and simulation.

Following the test run, samples of the naphtha feed and the resulting reformate were collected for octane number evaluation. The Research Octane Number (RON) and Motor Octane Number (MON) were measured using a hand-held Octanometer capable of measuring both RON and MON. The instrument used for the measurements is shown in Plate 7.

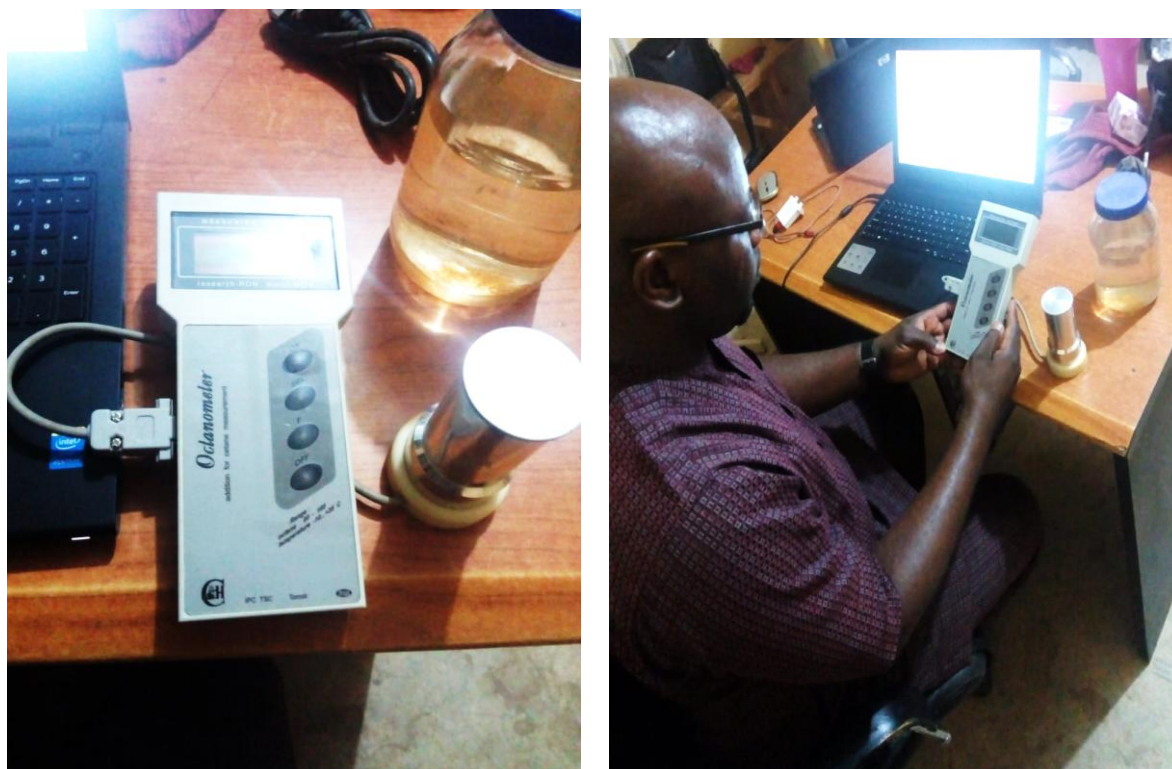


Plate 7: Hand-held Octanometer used for experimental determination of RON and MON of the naphtha feed and reformate.

The feed naphtha recorded 62 RON and 58 MON, whereas the reformate obtained from the fabricated catalytic reforming unit recorded 71 RON and 64 MON, respectively. Thus, the experimental process resulted in an increase of 9 RON points and 6 MON points relative to the feedstock. The experimental results are presented in Table 7.

Table 7: RON and MON of the feed and reformate

Parameter	Feed	Experimental Reformate	Predicted Reformate
RON	62	71	100
MON	58	64	84

The experimental results from Table 7 demonstrated that the fabricated reforming unit produced a measurable improvement in the octane quality of the heavy naphtha feed. However, the experimentally obtained RON of 71 and MON of 64 values were lower than the predicted reformate values (RON of 100 and MON of 84) and the higher octane values generally associated with commercial catalytic reforming. This difference is attributable to the disparity between the design simulation conditions and the actual test conditions as presented in Table 6. In particular, the experimental run was conducted at approximately 3 bar and without external hydrogen addition, while hydrogen addition was included in the design and simulation basis. The actual operating conditions, together with the use of manually operated controls and possible limitations in catalyst performance, may therefore have contributed to the lower experimental octane improvement.

The available experimental record contains single reported measurements for the feed and reformate. Replicate measurements and associated experimental uncertainty were not recorded during the original test campaign. Consequently, statistical uncertainty is not reported for the RON and MON values. Future experimental studies should include replicate measurements, documented instrument calibration and measurement uncertainty to improve the reproducibility and statistical reliability of the octane results.

Table 8: Comparison of the fabricated pilot-scale catalytic reforming unit with established industrial practice

Parameter	Fabricated pilot-scale unit	Established industrial practice
Capacity	5 BSPD	Published industrial catalytic reformers include 32,000 BSPD (Paulsboro); 40,000 BSPD (Chalmette); 35,000 BSPD (KNPC) reforming units.
Reactor configuration	3 fixed-bed reactors	3–4 reactors in series are used in conventional fixed-bed/semi-regenerative catalytic reforming configurations.
Catalyst	Pt–Re/Al ₂ O ₃	Pt–Re/Al ₂ O ₃ is an established catalyst formulation for naphtha reforming.
Reformate RON	71	Approximately 95–102, depending on feedstock and reforming severity.

Source: Present study; Yusuf et al. (2019); Mazziere et al. (2005); U.S. Energy Information Administration (EIA) (2026), Kuwait National Petroleum Company (KNPC) (2025).

The 5 BSPD capacity of the fabricated plant is quite considerably smaller as compared to that of existing commercial industrial catalytic reforming plants (See Table 8). Existing operational commercial plants include 32,000 BSPD catalytic reforming plant of the Paulsboro refinery and 40,000 BSPD catalytic reforming plant of the Chalmette refinery. Moreover, there are two continuous catalytic reforming plants of capacity 35,000 BPD installed in Kuwait National Petroleum Company (KNPC) Clean Fuels Project. This is just an illustration of the difference in the capacity of the local fabricated pilot plant as opposed to commercial refinery reformers.

4. CONCLUSIONS AND RECOMMENDATIONS

4.1 Conclusions

- This study successfully achieved the design, fabrication, installation and test-run of a 5 BSPD pilot-scale semi-regenerative catalytic naphtha reforming unit for the processing of straight-run heavy naphtha. The study commenced with characterization of the Escravos heavy hydrotreated naphtha used as the design feedstock and establishment of the process design basis. The selected semi-regenerative configuration comprised three fixed-bed catalytic reactors arranged in series, with interstage heating, consistent with the requirements of catalytic naphtha reforming.
- The process material balance established a design feed rate of 25.72 kg/h to the reforming section. The component balances demonstrated progressive conversion of paraffins and naphthenes to aromatics through the three reactor stages, accompanied by hydrogen generation. The paraffin content decreased from 8.57 kg/h in the feed to 0.55 kg/h at the outlet of R-03, while naphthenes decreased from 10.54 to 2.52 kg/h. In contrast, aromatics increased from 6.51 to 14.53 kg/h and hydrogen increased from 0.10 to 8.12 kg/h. The final separator further divided the reactor effluent into an H₂-rich gas stream and liquid reformate, demonstrating the intended material-separation function of the process.
- The energy-balance calculations provided the thermal design basis for the process equipment. The calculated heat-transfer duty for EX-01 was 74.62 MJ/h, while the fired-heater duties for H-01, H-02 and H-03 were 201.58, 1088.55 and 11,800.65 MJ/h, respectively. The corresponding calculated endothermic heat requirements of R-01, R-02 and R-03 were 8,289.82, 11,787.37 and 11,797.63 MJ/h, respectively. These calculations, together with the material balances, provided the basis for sizing and specifying the principal equipment of the pilot plant, including the surge drum, heat exchanger, fired heaters, fixed-bed reactors and phase separator.
- The resulting equipment was successfully fabricated and assembled into an operational pilot-scale reforming unit. The fabricated system comprised surge drum D-01, feed/product heat exchanger EX-01, phase separator D-02, three fired heaters (H-01, H-02 and H-03) and three fixed-bed

- reactors (R-01, R-02 and R-03). The successful installation and test run demonstrate the practical feasibility of translating the process design and equipment calculations into an indigenous pilot-scale catalytic reforming facility.
- v. Experimental evaluation of the produced reformat gave a RON of 71 and MON of 64, compared with feed values of 62 and 58, respectively. Thus, the reforming operation produced an improvement in both measured octane indices, although the resulting reformat quality remained below the commercial gasoline quality values considered in this study. The relatively modest octane improvement is interpreted in the context of the actual pilot-plant operating conditions, including the relatively low operating pressure, operation at approximately 460–480 °C, absence of hydrogen addition during the test run and manual process control. Other process variables may also have contributed to the observed product quality. The available experimental results do not permit the effect of any single factor to be isolated.
 - vi. The study demonstrates the feasibility of designing and fabricating an indigenous 5 BPSD catalytic naphtha reforming pilot unit and provides a practical platform for further investigation of catalyst performance, process optimization, reformat quality and hydrogen production. The fabricated unit also provides an experimental bridge between process-design calculations and practical refinery-process operation and can support subsequent studies aimed at optimizing the operating conditions and improving the performance of locally developed refining technology.

4.2 Recommendations

The following recommendations can be made based on the limitations and results obtained from this study:

- i. Operating conditions should be optimized in the fabricated pilot-scale catalytic reforming unit with particular emphasis being put on reactor temperature, operating pressure, hydrogen to hydrocarbon ratio and feed rate to enhance the severity of the reforming process and improve the quality of octane in the produced reformat.
- ii. Hydrogen circulation or hydrogen to hydrocarbon ratio should be applied in future experiments in order to investigate its effect on catalyst stability, reforming performance and quality of RON and MON in the produced reformat.
- iii. The pilot plant should be installed with better instruments and process-control system for improved and better controlled reforming process with respect to reactor temperature, pressure, feed rate and other operating conditions.
- iv. Further experiments on the fabricated pilot unit should be done at different operating conditions for establishment of process variables versus reformat octane quality and identification of optimum operating conditions for the unit.

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