DESIGN AND FABRICATION OF A FLUID CATALYTIC CRACKING UNIT OF A MINI REFINERY

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BY

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A DISSERTATION SUBMITTED TO THE SCHOOL OF POSTGRADUATE STUDIES, AHMADU BELLO UNIVERSITY, ZARIA IN PARTIAL FULFILLMENT OF THE REQUIREMENTS FOR THE AWARD OF MASTER OF SCIENCE DEGREE IN CHEMICAL ENGINEERING

DEPARTMENT OF CHEMICAL ENGINEERING,

FACULTY OF ENGINEERING

AHMADU BELLO UNIVERSITY, ZARIA

NIGERIA

APRIL, 2016

Declaration

I declare that the work in this Thesis titled 'Design and Fabrication of a Fluid Catalytic Cracking Unit of a Mini-Refinery' has been carried out by me in the Department of Chemical Engineering. The information derived from the literature has been duly acknowledged in the text and a list of references provided. To the best of my knowledge no part of this thesis was previously presented for another degree or diploma at this or any other institution.

_____ _____

Shehu, Mukhtar Date

Certification

This Thesis entitled DESIGN AND FABRICATION OF A FLUID CATALYTIC CRACKING UNIT OF A MINI-REFINERY by Mukhtar SHEHU meets the regulations governing the award of Master of Science Degree of the Ahmadu Bello University, and is approved for its contribution to knowledge and literary presentation.

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Dedication

I dedicate this work to my late father – Malam Shehu Salihu and to my beloved mother – Hajiya Aishatu Abubakar.

Abstract

Nigeria is one of the top oil-producing nations in the world but processing this oil into finished products has been a major challenge for the countries. Fluid Catalytic Cracking (FCC) Units of Nigerian Refineries produced petroleum products at far below their installed capacities.This work is aimed at effective design and fabrication of a mini FCC Unit.

The FCCU was designed to process 0.56bbl/hr of Escravos gas oil and fabrication of same was carried out. Equipments fabricated include Feed Surge Drum (Diameter 0.631m and Height 1.002m), Feed/LCO Heat Exchanger (Shell Diameter 0.260m, Tube Diameter 0.030m and Exchanger Length 1.000m), Feed/DCO Heat Exchanger(Shell Diameter 0.275m, Tube Diameter 0.0245m and Exchanger Length 1.000m) , Feed/Slurry Heat Exchanger(Shell Diameter 0.260m, Tube Diameter 0.024m and Exchanger Length 1.000m) , Feed Fired Heater (Base Diameter 0.872m, Height 1.745m and Tube Diameter 0.024m), Riser Reactor (Diameter 0.545m and Height 1.817m), Main Fractionator (Column Diameter 0.520m and Column Height 2.925m) and Overhead Condenser (Shell Diameter 0.310m, Tube Diameter 0.0245m and Condenser Length 1.000m).

Both manual and Hysys simulation software were used to carryout the design and the results obtained were compared to ascertain most acceptable technique for the design of FCCU.While Hysys simulation gave more detailed and effective results, the manual design technique shows little descrepancy to Hysys simulation result. Therefore, the manual design technique is also valid and can also serve in the absence of the software for FCCU design.

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Symbols and abbreviations

FCC Fluid Catalytic Cracking

LCO Light Cycle Oil

HCO Heavy Cycle Oil

DCO Decanted Oil

VGO Vacuum Gas Oil

HGO Heavy Gas Oil

PV Plug Valve

SLV Slide Valve

MAT Micro-activity testing

ACE Advance Catalyst Evaluation

PONA Paraffinic, Olefinic, Naphthenic and Aromatics

ASTM American Society for testing and Materials

TBP True Boiling Point (^{o}C)

IBP Initial Boiling Point (^{o}C)

ETP End Boiling Point (^{o}C)

SUV Saybolt Universal Viscometer

SFV Saybolt Flural Viscometer

ER&E Exxon Research and Engineering

KBR Kellogg Brown & Root

SWEC Stone & Webster Engineering Corporation

UOP Universal Oil Products

TDH Transport Disengaging Height (m)

KRPC Kaduna Refining and Petrochemical Company

WRPC Warri Refinery and Petrochemical Company

PHRC Port Harcourt Refinery Company Limited PHRC

FBR	Fluidized Bed Reactor
R	Bed Expansion Ratio
r	Bed Fluctuation Ratio
FQ	Fluidization quality
u_{mf}	Minimum Fluidization Velocity Bed (m/s)
ΔP	Pressure Drop (atm)
Ws	Mass of solids in the bed (kg)
P_s	Density of solid (kg/m^3)
A_s	Cross-sectional area of solid (m^2)
h	Height of the bed settled before the particles start to lift (m)
ϵ	Void fraction of bed
μ	Fluid viscosity (kg/ms)
ρ_f	Fluid density (kg/m^3)
u	fluid velocity (m/s)
d_p	Particle diameter (micron)
U_{mf}	Minimum fluidization velocity (m/s)
g	Acceleration due to gravity (m^2/s)
ρ_f	Fluid density (kg/m^3)
ρ_d	Particle density (kg/m^3)
ϵ_{mf}	Void fraction of bed at minimum fluidization

CHAPTER ONE

INTRODUCTION

1.1 Background to the Study

Petroleum refinery is a complex industry that generates a diverse slate of fuel and chemical products, from gasoline to heating oil (Rader, 1996). The refining process involves separating, cracking, restructuring, treating, and blending hydrocarbon molecules to generate petroleum products. Technological perspective is essential for a basic understanding of the complex refinery processes, a design based perspective is essential to develop a greater insight with respect to the physics of various processes, as design based evaluation procedures enable a successful correlation between fixed and operating costs and associated profits.

A refinery is a chemical plant that processes crude oil and produces several valuable products; it contains different types of units that perform a variety of different operations. The main goal is to take the undesirable components of the crude oil and upgrade them into more valuable products. Gasoline, diesel, and jet fuel are among the most valuable products. Refineries perform three basic operations which are Separation (fractional distillation), Conversion (cracking and rearranging the molecules), and Treatment.

Fluid Catalytic Cracking process is an important and widely used way to convert heavy feedstock into lighter, more valuable products. There are approximately 400 FCC units operating worldwide, with total processing capacity of over twelve million barrels per day (12 MMbbl/day) (Hug, 1998). Various feedstocks can be used, such as gas oils, vacuum gas oils or residual materials. Typical products are gasoline, light fuel oils and olefin-rich gases. The principal purpose of a cracking unit is to break high molecular weight hydrocarbons into smaller pieces of

lower boiling point fractions, especially gasoline (Dwyer and Rawlence, 1993). Originally, thermal operations were used to crack heavy oil, but the discovery of a catalyst that gives a higher yield of gasoline with a higher octane number quickly brought on the use of catalytic cracking units. Today, the most commonly used catalytic cracking unit is the Fluid Catalytic Cracker or FCC (Wilczura-Wachnik, 1973). The fluid cracker consists of a catalyst section and a fractionating section that operate together as an integrated processing unit. The catalyst section contains the reactor and regenerator, which, with the standpipe and riser, forms the catalyst circulation unit (Ibsen, 2006).

This research work is intended at the design and fabrication of a fluid catalytic cracking unit of a Mini-Refinery for the ultimate purpose of improving present yield of gasoline in Nigerian refineries through pilot testing in the mini refinery FCCU of improved catalyst and different feed composition.

1.2 Problem Statement

Nigeria is one of the top oil-producing nations in the world but processing this oil into finished products has been a major challenge for the country. The Nigerian FCCU produced petroleum products at far below their installed capacity and this is as a result of neglect by stake holders in areas of research and developments in the fields of enhancing catalyst development and new FCC feedstock. Also this due to unavailability of operational data that can be used to improve the production capacity. There is therefore the need to provide testing base for research in both fields and also in providing operational data to boost oil processing capacity and be self-reliant when it comes to petroleum and petroleum processing.

1.3 Justification

i- Pilot plants can serve as small scale of larger commercial units.

ii- Fabricated Pilot FCCU can serve as study aid for students of chemical engineering and related fields.

iii- Research and development data will be provided for specific FCC feed stock.

iv- Research into catalyst development & testing will be enhanced.

v- Job creation

1.4 Aim and Objectives

The aim of this work is to design and fabricate a 5 Barrels per batch Mini Fluid Catalytic Cracking Unit.

The specific objectives of this work include:

i- Identify suitable process selection for the Pilot plant

ii- Carry out material and energy balances

iii- Development of a Preliminary design of equipments

iv- Detailed design of Major equipments

v- Plant layout

vi- Instrumentation and control

vii- Fabrication of the Pilot Plant.

1.5 Scope

The scope of this work is limited to the design and fabrication of a Fluid Catalytic Cracking Unit of a Mini – Refinery to process 5 barrels per batch in Nigeria as well as the safety consideration of such a plant.

CHAPTER TWO

LITERATURE REVIEW

2.1 Historical Background of Petroleum Refining

Petroleum refining has evolved continuously in response to changing consumer demand for better and different products. The original requirement was to produce kerosene as a cheaper and better source of light than whale oil. The idea of cracking large hydrocarbon molecules into smaller ones began in the late 1800's when it was found that the large molecules broke into smaller molecules at high temperatures (Rader, 1996). There are several factors that complicate FCC operation, among them feed composition is probably the main one, since it is a complex mixture of hydrocarbons with different chemical reactivity (Mariaca-Domínguez et al., 2003).

2.2 Crude Oil and its Constituents

Petroleum crude oil consists primarily of a mixture of hydrocarbons with small amounts of other organic compounds containing sulfur, nitrogen and oxygen. The crude oil also contains small amounts of metals such as copper, iron, nickel and vanadium. Hundreds of crude oils (usually identified by geographic origin) are processed, in greater or lesser volumes, in the world's refineries. Each crude oil is a unique mixture of thousands of compounds, mainly hydrocarbons (Rader, 1996) .

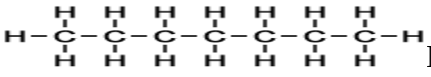
2.2.1 Hydrocarbon classification

The hydrocarbon types in the FCC feed are broadly classified as Paraffinic, Olefinic, Naphthenic and Aromatics (PONA)

2.2.2 Hydrocarbon compounds

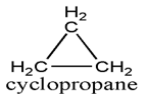
1. Alkanes or Paraffins:

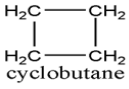
Alkanes are saturated compounds having the general formula C_nH_{2n+2}. Alkanes are relatively nonreactive compounds in comparison to other series. They may either be straight-chain or branched-chain compounds.

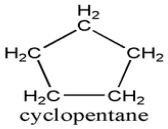

N-Heptane

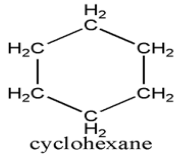
2. Cycloalkanes or Cycloparaffins (Naphthenes):

Cycloalkanes and bicycloalkanes are normally present in crude oils and its fractions in variable proportions. The presence of large amounts of these cyclic compounds in the naphtha range has its significance in the production of aromatic compounds. Naphtha cuts with a high percentage of Naphthenes would make an excellent feedstock for aromatization.


cyclopropane

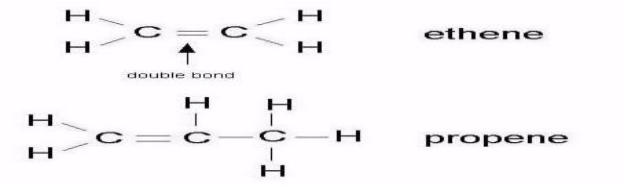

cyclobutane


cyclopentane


cyclohexane

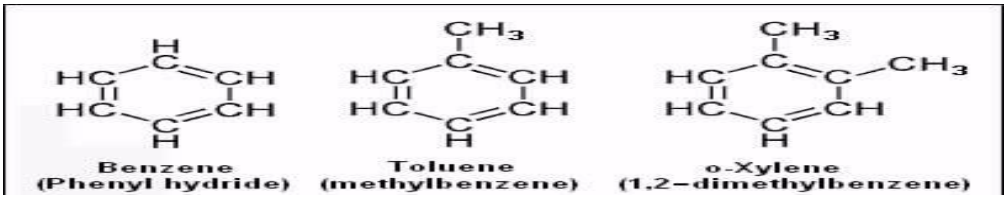
3. Alkenes or Olefins:

Alkenes are unsaturated hydrocarbon compounds having the general formula CnHn. They are practically not present in crude oils, but they are produced during processing of crude oils at high temperatures.



4. Aromatics:

Aromatic compounds are normally present in crude oils. Only monomolecular compounds in the range of C6–C8 (known as B-T-X) have gained commercial importance.



2.2.3 Related works

Related works reviewed are presented in Table 2.1:

Table 2.1: Review of related work

S/N	Author	Title of Work	Conclusion
1.	Leuenberger *et al.*, 1988	Pilot scale equipment testing for optimum performance	Pilot Plant Offers advantages and fewer risks than laboratory equipment to scale up operating conditions and process performance, quickly generating enough quantities of products for more complete and detailed analysis.
2.	Boock & Zhao,1998	A trend to develop better catalysts in order to look for optimal operating conditions.	Due to FCC's high impact on the profitability of refineries, it is often unfeasible" and/or impossible to try new feedstock or catalysts during FCC industrial operation.

3.	Maya-Yescas et al. , 2006	Fluidized bed catalytic cracking units emulation in pilot plant	Laboratory micro activity plants, like Micro-activity testing (MAT) and Advance Catalyst Evaluation (ACE) units, give quick results that are not usable, directly, to model the industrial process performance.
4.	Latridis et al., 2010	FCC pilot plant reactor design for improved yields	It is shown that for achieving product yields consistent with the range of commercial FCC unit operating conditions, a minimum of riser length is required.
5.	Bielansky et al., 2010	Processing of pure vegetable oils in a continuous FCC pilot plant.	The addition of vegetable oil had hardly any influence on the cracking Process. The gasoline achieved is oxygen free at high octane numbers.
6.	Bryden, 2013	Predicting FCC unit performance with pilot plant testing	" In order to minimize risk, catalyst testing using a pilot plant is important in predicting commercial unit performance"

2.3 FCC Feed Characterization

Refiners process many different types of crude oil. As market conditions and crude quality fluctuate, so does catalytic cracking feedstock. Often the only constant in FCC operations is the continual change in feedstock quality (Leuenberger *et al*., 1988). Feed Characterization is the process of determining the physical and chemical properties of the feed. Two feed with similar boiling point ranges may exhibit dramatic differences in cracking performance and product yields. FCC feed characterization is one of the most important activities in monitoring catalytic cracking operation. Understanding feed properties and knowing their impact on unit performance

are essential. Troubleshooting, catalyst selection, unit optimization and subsequent process evaluation all depend on the feed stock (Reza, 2000).

Table 2.2 shows a widely-used scheme for classifying crude oils on the basis of their API gravity and sulfur content. Each crude class is defined by a range of API gravity and a range of sulfur content.

Table 2.2: Classification of different types of Crude oil

Crude Oil	**Country of Origin**	**Crude Oil Class**	**Gravity (°API)**	**Sulfur (wt. %)**
Brent	U.K.	Light Sweet	40.0	0.5
West Texas Intermediate	U.S.A.	Light Sweet	39.8	0.3
Arabian Extra Lt. Export	Saudi Arabia	Light Sour	38.1	1.1
Daqing	China	Medium Medium Sour	33.0	0.1
Forcados Export	Nigeria	Medium Medium Sour	29.5	0.2
Arabian Light Export	Saudi Arabia	Medium Sour	34.0	1.9
Kuwait Export Blend	Kuwait	Medium Sour	30.9	2.5
Marlim Export	Brazil	Heavy Sweet	20.1	0.7
Maya Heavy Export	Mexico	Heavy Sour	21.3	3.4

Source: Kaduna Refining and Petrochemical Company Handbook, 1980.

2.3.1 Feedstock physical properties

Characterizing an FCC feedstock involves determining both its chemical and physical properties because sophisticated and analytical techniques, such as mass spectrometry are not practical on a daily basis, physical properties are used. They provide qualitative measurement of the feed's composition (Leuenberger *et al*., 1988). The refinery laboratory is usually equipped to carry out these physical property tests on a routine basis. The most widely used properties are:

i- Density (kg/m^3)

ii- Watson Factor KW

iii- True Boiling Point (^oC)

iv- Pour Point (^oC)

v- Cloud Point (^oC)

vi- Octane Number

vii- Sulphur Content (ppm)

viii- Refractive Index

ix- Bromine Number and Bromine Index

x- Viscosity (kg/ms)

xi- Conradson, Rams bottom, Micro carbon, and Heptane Insolubles

2.3.2 Density

The main classification of crude is by referring to their density. Light crudes have a specific gravity (d) smaller than 0.825, medium crudes between 0.825 and 0.875, heavy crudes between 0.875 and 1.000 and extra heavy crudes larger than 1.000 (Reza, 2000)

2.3.2.1 API Gravity

While specific gravity may suffice, the petroleum industry uses another scale (API)

$$\text{API} = \frac{141.5}{d} - 131.5 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 2.1$$

Thus, light oils have a large API density and heavy ones a low one. Typical low values are close to 15 (some Venezuelan oils) to about 40 (some middle east oils). Crudes are also characterized as paraffinic, asphaltic, naphthenic and aromatic. This primary classification intends to reflect the relative abundance of different types of compounds, as their name indicates. These can also be more characterized in more detail by the density of their heavy gasoline cut (boiling point between 100 and 200 ^{o}C), d1, and the residue (boiling point larger than 350 ^{o}C), d2. For example a paraffinic crude has d1<0.76 and d2<0.93, an aromatic crude has d1>0.80 and d2 between 0.93 and 0.975, an asphaltic crude has d1>0.78 and d2>0.975, and so on (Reza, 2000). This is obviously related to the relative abundance of hydrogen and carbon. Nevertheless, there is a need for a unique number, rather than a range, to characterize the type of crude and this is somewhat reflected by the Watson factor (KW).

2.3.3 Watson factor KW

This factor, which was introduced by researchers at Universal Oil products Company

(UOP) and for this reason also known as (KUOP), is based on the observation that the H/C ratio is related to the boiling points of mixtures, as it is to the individual components.

Thus, for pure components, we have

$$\text{KW} = \frac{(1.8T)1/3}{d} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 2.2$$

Where (T) is the boiling point (Kelvin) and, d the specific gravity at 15.6 ^{o}C (60 ^{o}F). This value is typically 13 for Paraffins, 12 for hydrocarbons whose chain and ring weights are equivalent, 11 for pure Naphthenes and 10 for pure aromatics. Because mixtures do not have a unique boiling point, to use this factor in mixtures one needs to resort to the use of some average temperature. Thus, one needs to record the temperatures for which 10, 20, 50, 80 and 90% of the volume was distilled from the sample. Thus for TBP distillation (true boiling point), the following is used:

$$T = \frac{T20+T50+T80}{3} \quad \text{.....2.3}$$

For ASTM distillations, the following is used

$$T = \frac{T10+2T50+T90}{4} \quad \text{.....2.4}$$

While use of this factor has been extremely useful to characterize crudes, the latest tendency, albeit more expensive and time consuming, is to rely on better characterization methods based on modern physical tests, like Refractive index and Molecular weight Method, mass spectrometry, ultraviolet (UV) spectrometry, carbon atom distribution by infrared spectrometry and nuclear magnetic resonance, or by chromatographic (gas and liquid) techniques.

2.3.4 True boiling point (TBP) method for crude characterization

The true boiling point curve is basically a plot of the boiling point of each component of the mixture as a function of the cumulative volumetric fraction distilled. It implies that a batch distillation column with a large number of trays and a large reflux ratio is used, which will

ensure that each component is separated at a time. In practice, the ASTM D2892 test method (ASTM is an acronym for American Society for testing and Materials) is used. It consists of using a 15 to 18 tray distillation column operating at a 5:1 reflux ratio. The distilled volume is measured and the temperature in the reboiler is recorded (Reza, 2000).

2.3.5 Pour point:

Defined as the temperature at which the oil ceases to flow. The pour point is obviously related to pumping problems, which are obviously avoided by heating up the mixture.

2.3.6 Cloud point:

Cloud point is the temperature at which the oil becomes cloudy or hazy.

2.3.7 Octane number

Octane rating or octane number is a standard measure of the performance of a motor or aviation fuel. The higher the octane number, the more compression the fuel can withstand before detonating (igniting). In broad terms, fuels with a higher octane rating are used in high performance petrol engines that require larger compression ratios. In contrast, fuels with lower octane numbers (but higher cetane numbers) are ideal for diesel engines, because diesel engines (also referred to as compression-ignition engines) do not compress the fuel but rather compress only air and then inject the fuel into the air heated up by compression, whereas petrol engines (also referred to as gasoline engines) rely on ignition of air and fuel compressed together as a mixture without ignition, which is then ignited at the end of the compression stroke using spark plugs. Therefore, high compressibility of the fuel matters greatly for only petrol engines. Use of gasoline (also known as petrol) with lower octane numbers may lead to the problem of engine knocking.

2.3.8 Sulfur content:

The crude oil to be processed determine the amount of sulfur in the product hence different ways of predicting the sulfur content of products are available depending on the product cuts. FCC feedstock contain sulfur in the form of organic sulfur compound such as mercaptan,sulfide and thiophenes. Frequently as the residue content of crude oil increases so as the sulfur content (Reza, 2000) total sulfur in FCC feed is determined by the wavelength dispersive x-ray florescence spectrometry method (ASTM D-2622).

2.3.9 Refractive index

Refractive index (RI) measure how refractive or aromatic a sample is. The higher the RI, the more the aromatics and the less crackable the sample. A feed having an RI of 1.5105 is more difficult to crack than a feed with an RI of 1.4990 (Reza, 2000). The RI can be measured in a lab (ASTM D-1218) or can be predicted using carrelations.

2.3.10 Bromine number and bromine index

Bromine number (ASTM D-1159) and bromine index (ASTM D- 2710) are qualitative methods to measure the reactive sites of a sample (Reza, 2000). Bromine reacts not only with olefin bonds, but also with basic nitrogen molecules and with some aromatic sulfur derivatives. Nevertheless, olefins are the most common reactive sites and the bromine number used to indicate olefinicity of the feed. The bromine number is the number of grams of bromine that will react with 100 grams of the sample. Typical bromine numbers are:

i- Less than 5 for hydro treated feeds

ii- 10 for heavy vacuum gas oil

iii- 50 for Coker gas oil

A general rule-of-thumb is that the olefin fraction of the sample is one-half of its bromine number. Alternatively, the bromine index is the number of milligrams of bromine that will react with 100 grams of the sample, and is most often used by the chemical industry for stocks that have very low olefin content (Reza, 2000).

2.3.11 Viscosity

Viscosity indicates the chemical composition of an oil sample. As the viscosity of a sample increases, paraffins increase, hydrogen content increases, and the aromatic fraction decreases (Reza 2000). Viscosity is normally measured at two different temperatures: typically 100.4°F (38°C) and 210°F (99°C). For many FCC feeds, the sample is too thick to flow at 100°F and the sample usually is heated to about 130°F (Reza, 2000). Viscosity is a measurement of resistance to flow. Although the unit of absolute viscosity is poise, its measurement is difficult. Instead, kinematic (flowing) viscosity is determined by measuring the time for given flow through a capillary tube of specific diameter and length. The unit of kinematic viscosity is the Stoke. However, in general practice, centistokes is used. Saybolt viscosity (ASTM D-88) is the most popular method of measuring the viscosity of oils such as FCC feed. This method covers two procedures:

i- Saybolt Universal Viscometer (SUV) for light oils

ii- Saybolt Flural Viscometer for (SFV) for Heavy oils

Both procedures measure the time for fixed volume of sample to flow through a calibrated tube at a controlled temperature.

2.3.12 Conradson, ramsbottom, micro carbon, and heptane insolubles

One area of catalytic cracking not fully understood is the proper determination of carbon residue of the feed and how it affects the unit's coke formation on the catalyst. Carbon residue is defined as the carbonaceous residue formed after thermal destruction of a sample. Catalytic crackers are generally limited in coke burn capacity; therefore, the inclusion of residue in the feed produces more coke and forces a reduction in FCC throughput. Conventional gas oil feeds generally have a carbon residue less than 0.5 wt%; for feeds containing residue, the number can be as high as 15 wt%. Four popular tests are presently used to measure carbon residue or concarbon of FCC feedstocks:

i- Conradson

ii- Ramsbottom

iii- Micro-method

iv- Heptane Insolubles

The object is to indicate the relative coke forming tendency of feedstocks. Each test has advantages and disadvantages, but none of them provide a rigorous definition of carbon residue or asphaltenes.

The Conradson test (ASTM D-189) measures carbon residue by evaporative and destructive distillation. The sample is placed in a pre weighed sample dish. The sample is heated, using a gas burner, until vapor ceases to burn and no blue smoke is observed. After cooling, the sample dish is reweighed to calculate the percent carbon residue. The test, though popular, is not a good measure of the coke forming tendency of FCC feed because it indicates thermal, rather than catalytic, coke. In addition, the test is labor intensive and is usually not reproducible, and the procedure tends to be subjective (Reza, 2000).

The Rams bottom test (ASTM D-524) is also used to measure carbon residue. The test calls for introducing 4 grams of sample into a pre weighed glass bulb, then inserting the bulb in a heated bath for 20 minutes. The bath temperature is maintained at 1,027°F (553°C). After 20 minutes, the sample bulb is cooled and reweighed. Compared with the Conradson test, Rams bottom is more reproducible and precise.

The Micro-method uses an analytical instrument to measure Conradson carbon in a small automated set. The Micro method (ASTM D-4530) gives test result that equivalent to the Conradson carbon residue test (D 189). This test serve as an indication of the relative coke forming tendency of such material.

The heptane insoluble method (ASTM D-3279) is commonly used to measure the asphaltene content of the feed. Asphaltenes are insoluble in c_3 to c_7 paraffins. The amount of asphaltenes that precipitate varies from one solvent to another, so it is important that the reported asphaltene values be identified with the appropriate solvent. Both normal heptane and pentane insolubles are widely used for measuring asphaltenes

2.3.13 Impurities

The Catalytic cracker as the main conversion unit is designed to handle a variety of feed stocks. Today's FCC feedstocks are generally heavier and contain higher levels of impurities like nitrogen, Sulphur and metals. These impurities have negative effects on unit performance. Understanding the nature and effects of these contaminants is essential in feed and catalyst selection as well as troubleshooting the unit. Most of the impurities in the FCC feed exist as components of large organic molecules. The most common contaminants are:

i- Nitrogen

ii- Sulphur

iii- Nickel

iv- Vanadium

v- Sodium

vi- Copper

vii- Iron

With the exception of sulfur, all these conterminants poison FCC catalyst, causing it to lose its ability to produce valuable products. Sulfur on the other hand increases operating costs because additional feed and product treatment facilities are required to meet product specifications and compliance with environmental regulations.

Nitrogen in FCC feed refers to organic nitrogen compounds. The nitrogen content of FCC feed is usually reported as basic and total nitrogen (Reza 2000). Basic nitrogen will neutralize acid sites on the catalyst. This results to temporary loss of catalyst activity and a drop in unit cinversion. However, the poisoning is temporary because burning of nitrogen in the regenerator restores the activity of the catalyst. About 70% to 90% of the nitrogen in the coke is coverted to elemental nitrogen while the remaining nitrogen is converted to nitrogen oxides (NO_x) which leaves the unit with the flue gas. Table 2.3 presents a typical properties of the feed to the KRPC FCCU. From the table it can be observed that the feed are of three different properties for normal FCC, Resid FCC and Atmaspheric residue.

Table 2.3 Typical KRPC Feed Properties

	FCC	**RFCC**	**AR**
S.G 15/4^{0}C	0.890	0.902	0.986
Nitrogen ppm	590	1840	1680
Viscosity Cst	190	160	146
Sulphur wt%	0.1	0.12	0.6
CCR wt%	1	4.7	5.9
Asphaltene wt%	0	0.1	3.1
V wt ppm	0	0	12
Ni wt ppm	0.2	4.4	8
DISTILLATION			
IBP ^{0}C	284	264	292
10%	351	296	383
30%	380	471	440
50%	409	543	495
70%	445	-	-
90%	490	-	-
K factor	12	12.5	11.9
Mol wt	371	580	466

Source: KRPC Handbook 1980.

2.3.14 Process variables

i- Temperature

ii- Pressure

iii- Catalyst to Oil ratio (Weight of catalyst per hour to weight of oil charged per hour)

iv- Space Velocity (Weight or volume of catalyst charged per hour/weight or volume of catalyst in the reaction zone per hour)

Wide flexibility in product distribution and quality is possible through the control of the process variable along with extent of internal recycle i.e heavy naphtha and heavy cycle oil pump around and refluxes (Hug,1998). Increased conversion can be obtained by higher temperature higher pressure, lower space velocity and higher catalyst to oil ratio.

2.3. 15.1 Dependent and independent variables

The Independent variables are Fresh feed rate, Feed Pre-heat temperature, Fresh catalyst activity/selectivity and Mode of combustion. While the Dependent variables are Regeneration Temperature, Regeneration Air rate and Conversion which can either be base on Reactor Temperature or Catalyst Circulation rate

2.4 FCC Catalyst

The introduction of Zeolite in commercial FCC catalyst in the early 1960s is one of the most significant advances in the history of Catalytic cracking. Zeolite catalyst provides a greater profit with little capital investment (Ibsen, 2006). Simply stated, Zeolite catalysts were and are still the biggest bargain of all time for the refiner. Improvements in catalyst technology have continued,

enabling refiners to meet the demands of their market with relatively minimum capital investment.

Compared to amorphous silica-alumina catalysts, the zeolite catalysts are more active and more selective, the higher activity and selectivity translate to more profitable liquid product yields and additional cracking capacity.

Modern FCC catalysts are fine powders with a bulk density of 0.80 to 0.96 g/cm3 and having a particle size distribution ranging from 10 to 150 μm and an average particle size of 60 to 100 μm (Reza 2000). The design and operation of an FCC unit is largely dependent upon the chemical and physical properties of the catalyst. The desirable properties of an FCC catalyst are:

i- Good stability to high temperature and to steam

ii- High activity

iii- Large pore sizes

iv- Good resistance to attrition

v- Low coke production

2.4.1 Catalyst component

A modern FCC catalyst has four major components: crystalline zeolite, matrix, binder, and filler. Zeolite is the primary active component and can range from about 15 to 50 weight percent of the catalyst (Dwyer and Rawlence, 1993). The zeolite used in FCC catalysts is referred to as faujasite or as Type Y and is composed of silica and alumina tetrahedra with each tetrahedron having either aluminum or silicon atom at the center and four oxygen atoms at the corners. It is a molecular sieve with a distinctive lattice structure that allows only a certain size range of hydrocarbon molecules to enter the lattice. In general, the zeolite does not allow hydrocarbon

molecules larger than 8 to 10 μm (i.e., 80 to 90 Angstroms) to enter the lattice (Dwyer and Rawlence, 1993).

The catalytic sites in the zeolite are strong acids (equivalent to 90% sulfuric acid) and provide most of the catalytic activity. The acidic sites are provided by the alumina tetrahedra. The aluminum atom at the center of each alumina tetrahedra is at a +3 oxidation state surrounded by four oxygen atoms at the corners which are shared by the neighboring tetrahedra. Thus, the net charge of the alumina tetrahedra is -1 which is balanced by a sodium ion during the production of the catalyst. The sodium ion is later replaced by an ammonium ion, which is vaporized when the catalyst is subsequently dried, resulting in the formation of Lewis and Bronsted acidic sites (Dwyer and Rawlence, 1993 and Reza, 2000). In some FCC catalysts, the Bronsted sites may be later replaced by rare earth metals such as cerium and lanthanum to provide alternative activity and stability levels.

The matrix component of an FCC catalyst contains amorphous alumina which also provides catalytic activity sites and in larger pores that allow entry for larger molecules than does the zeolite. That enables the cracking of higher-boiling, larger feedstock molecules than are cracked by the zeolite.

The binder and filler components provide the physical strength and integrity of the catalyst. The binder is usually silica sol and the filler is usually clay (kaolin) (Reza, 2000).

2.4.2 Zeolite chemistry

A typical zeolite consists of silicon and aluminum atoms that are tetrahedrally joined by four oxygen atoms. Silicon is in a +4 oxidation state; therefore, a tetrahedron containing silicon is

neutral in charge. In contrast, aluminum is in a +3 oxidation state. This indicates that each tetrahedron containing aluminum has a net charge of -1, which must be balanced by a positive ion. Solutions containing sodium hydroxide are commonly used in synthesizing the zeolite. The sodium serves as the positive ion to balance the negative charge of aluminum tetrahedron. This zeolite is called soda Y or NaY. The NaY zeolite is not hydrothermally stable because of the high sodium content. The ammonium ion is frequently used to displace sodium. Upon drying the zeolite, ammonia is vaporized. The resulting acid sites are both the Bronsted and Lewis types. The Bronsted acid sites can be further exchanged with rare earth material, such as cerium and lanthanum to enhance their strengths. The zeolite activity comes from these acid sites (Dwyer and Rawlence, 1993).

2.4.3 Zeolite types

Zeolites employed in the manufacture of the FCC catalyst are synthetic versions of naturally occurring zeolite called faujasites. There are about 40 known natural zeolites and over 150 zeolites that have been synthesized. Of this number, only a few have found commercial applications. The zeolites with applications to FCC are Type X, Type Y, and ZSM-5. Both X and Y zeolites have essentially the same crystalline structure. The X zeolite has a lower silica alumina ratio than the Y zeolite. The X zeolite also has a lower thermal and hydrothermal stability than the Y zeolite. Some of the earlier FCC zeolite catalysts contained X zeolite; however, virtually all of today's catalysts contain Y zeolite or variations thereof. ZSM-5 is a versatile zeolite that increases olefin yields and octane (Dwyer and Rawlence, 1993).

The zeolite catalyst has sites which can remove hydrogen from an alkane together with the two electrons which bound it to the carbon. That leaves the carbon atom with a positive charge. Ions

like this are called carbonium ions (or carbonations). Reorganization of these leads to the various products of the reaction.

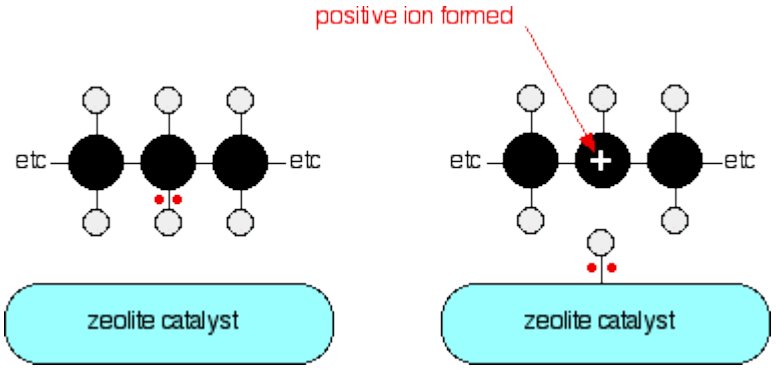


Figure 2.1 Zeolite Catalyst sites

2.4.4 Role of zeolite in FCCU

Synthetic zeolites are widely used as catalysts in Fluid catalytic cracking; Zeolites confine molecules in small spaces, which cause changes in their structure and reactivity. The hydrogen form of zeolites (prepared by ion-exchange) is powerful solid-state acids, and can facilitate a host of acid-catalyzed reactions, such as Isomerization, alkylation, and cracking. The specific activation modality of most zeolitic catalysts used in petrochemical applications involves quantum-chemical Lewis acid site reactions.

2.5 FCCU Pilot Plant

An important consideration in developing process technology applicable to commercial Fluid Catalytic Cracking Units (FCCU) is the operation of a pilot plant over a range of conditions of relevance to commercial operations. Typical C/O ratios in commercial FCC units are in the range of 4 to 9 (Latridis et al., 2010).

Fluidized Catalytic Cracking (FCC) is a very complex process combining a reaction stage (riser) with a continuous catalyst regeneration stage. Both stages strongly interact, particularly in energy balance.

Catalyst regeneration severity and capacity highly influences riser behaviour. There are several factors that complicate FCC operation, among them feed composition is probably the main one, since it is a complex mixture of hydrocarbons with different chemical reactivity.

Pilot plant studies are a support tool in research and development activities, as well as the platform for providing technical services and improvements. These plants are strategic bases for deciding about feedstock and catalyst, through evaluation of yields and quality of products. They are also important studies in feasibility. For example, in refining petroleum industry, and particularly in FCC (Maya-Yescas et al., 2006).

Pilot plant size allows studying and simulating confidently the industrial unit behaviour with no risk for production. It is also possible to develop new catalysts formulations for each feedstock in long term tests and give recommendations for better operation aiming to optimize conditions and improve profitability. Besides, pilot plant information is essential for developing, adjusting and validating FCC simulation models. As data source for simulation, pilot plants give opportunities to explore broad ranges of process conditions, being useful to process research and development (Maya-Yescas et al., 2006)

2.5.1 FCC Pilot plant description

The typical arrangement of a pilot plant scheme is shown in Fig. 2.2. Feedstock is taken from a couple of storage vessels, connected to the control system. This couple of storage vessels allows normalising operation with a reference feedstock and then introducing a new one. A dosage pump with precision control is used to send the feed through the heater and nozzle. There is a system to control and register the flow. A nitrogen or vapour stream can be used as dispersant,

feeding it through an independent heater. Feed vaporizes as soon as gets in touch with catalyst coming from regenerator and goes through the riser, as in the industrial unit. At the top of the riser, catalyst and products are separated using cyclones. Spent catalyst falls to a vertical column within a dense phase fluidized bed, in order to strip hydrocarbons from catalyst with vapour or nitrogen flowing counter current. A slide valve controls the flow of stripped catalyst to regenerator. Stripping temperature, bed height and vapour/nitrogen flows are controlled to adjust stripper efficiency. Gaseous products go to a stabilizer column, to separate components heavier than C5 & C6; this liquid product is fractionated to obtain gasoline, LCO and HCO (Maya-Yescas et al., 2006).

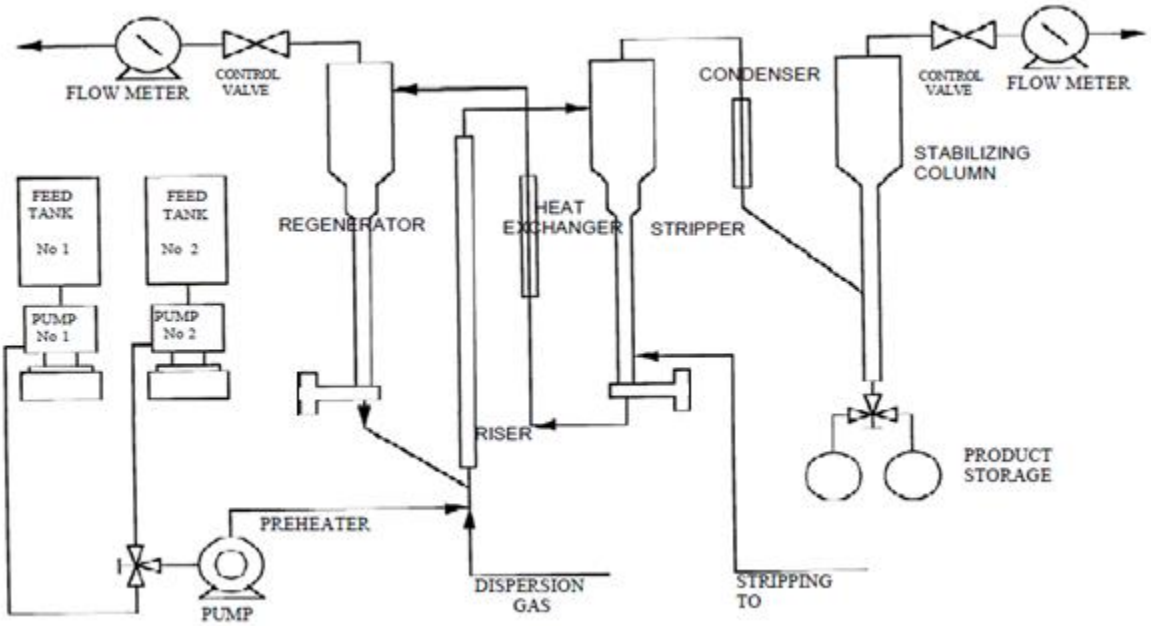


Figure 2.2: FCC Pilot Plant Basic Equipment

2.6 Cracking

Cracking is the name given to breaking up large hydrocarbon molecules into smaller and more useful bits. This is achieved by using high pressures and temperatures without a catalyst, or lower temperatures and pressures in the presence of a catalyst.

There isn't any single unique reaction happening in the cracker. The hydrocarbon molecules are broken up in a fairly random way to produce mixtures of smaller hydrocarbons, some of which have carbon-carbon double bonds. One possible reaction involving the hydrocarbon $C_{15}H_{32}$ might be:

$$C_{15}H_{32} \longrightarrow 2C_2H_4 + C_3H_6 + C_8H_{18} \quad \text{.....2.4}$$

ethene propene octane

Or, showing more clearly what happens to the various atoms and bonds:

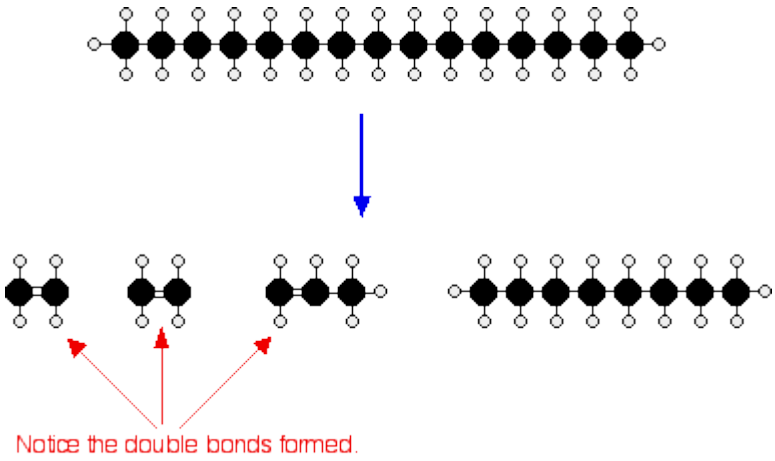


Figure 2.3 Cracking Reactions

This is only one way in which this particular molecule might break up. The ethene and propene are important materials for making plastics or producing other organic chemicals. The octane is one of the molecules found in petrol (gasoline).

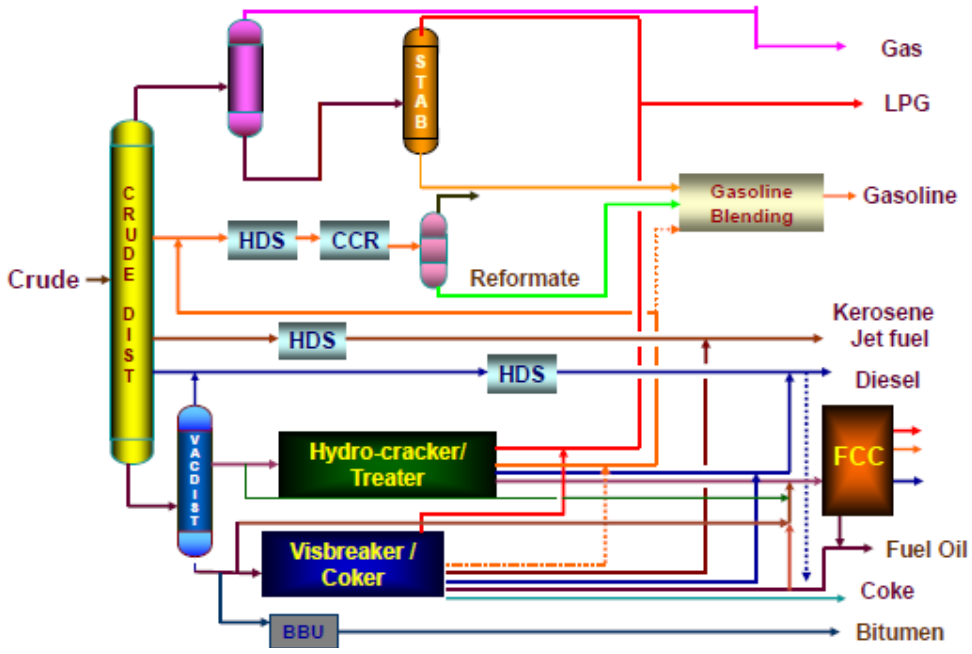
2.6.1 Fluid catalytic cracking

Modern cracking uses zeolite as the catalyst. These are complex aluminosilicates, and are large lattices of aluminum, silicon and oxygen atoms carrying a negative charge. They are, of course, associated with positive ions such as sodium ions.

The alkane is brought into contact with the catalyst at a temperature of about 500°C and moderately low pressures. The zeolites used in catalytic cracking are chosen to give high percentages of hydrocarbons with between 5 and 10 carbon atoms - particularly useful for petrol (gasoline). It also produces high proportions of branched Alkanes and aromatic hydrocarbons like benzene.

2.6.2 Fluid catalytic cracking unit

Fluid catalytic cracking (FCC) is one of the most important conversion processes used in petroleum refineries. It is widely used to convert the high-boiling, high-molecular weight hydrocarbon fractions of petroleum crude oils to more valuable gasoline, olefinic gases, and other products. Cracking of petroleum hydrocarbons was originally done by thermal cracking, which has been almost completely replaced by catalytic cracking because the latter produces more gasoline with a higher octane rating (Hug, 1998). It also produces byproduct gases that are more olefinic, and hence more valuable, than those produced by thermal cracking. Figure 2.4a shows the position Fluid Catalytic Cracking Unit in a Refinery Setting.

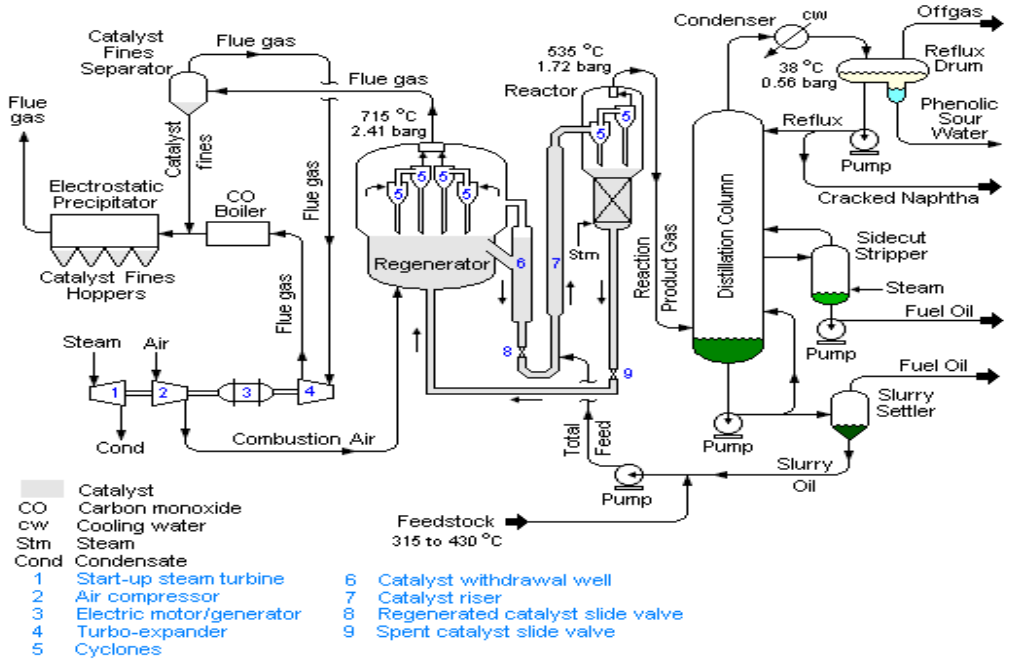


Source: (Bhattacharyya, 2013)

Figure 2.4a: Position of FCC in the Refinery

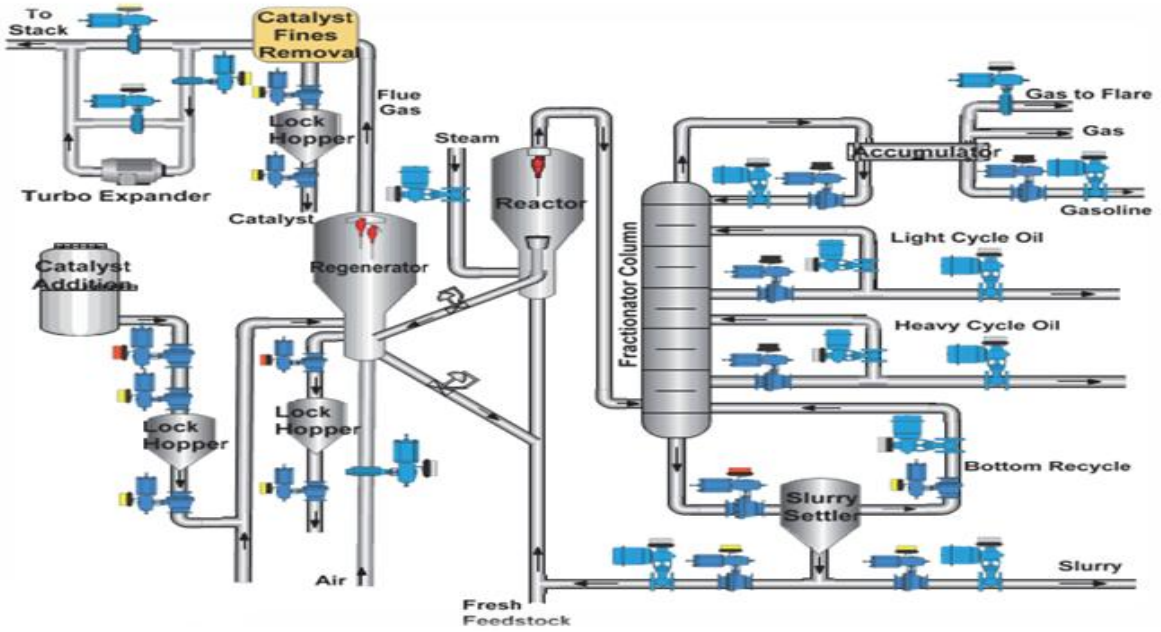
The feedstock to FCC is usually that portion of the crude oil that has an initial boiling point of 340 °C or higher at atmospheric pressure and an average molecular weight ranging from about 200 to 600 or higher (Reza, 2000). This portion of crude oil is often referred to as heavy gas oil (HGO) or vacuum gas oil (VGO). The FCC process vaporizes and breaks the long-chain molecules of the high-boiling hydrocarbon liquids into much shorter molecules by contacting the feedstock, at high temperature and moderate pressure, with a fluidized powdered catalyst (Figure 2.4b).

In effect, refineries use fluid catalytic cracking to correct the imbalance between the market demand for gasoline and the excess of heavy, high boiling range products resulting from the distillation of crude oil. A schematic diagram of a fluid catalytic cracking process is shown in figures 2.4b to 2.5. Figure 2.4b depict a side by side process showing process flows and conditions while 2.8 shows automations and 2.9 shows a schematic presentation of the process.



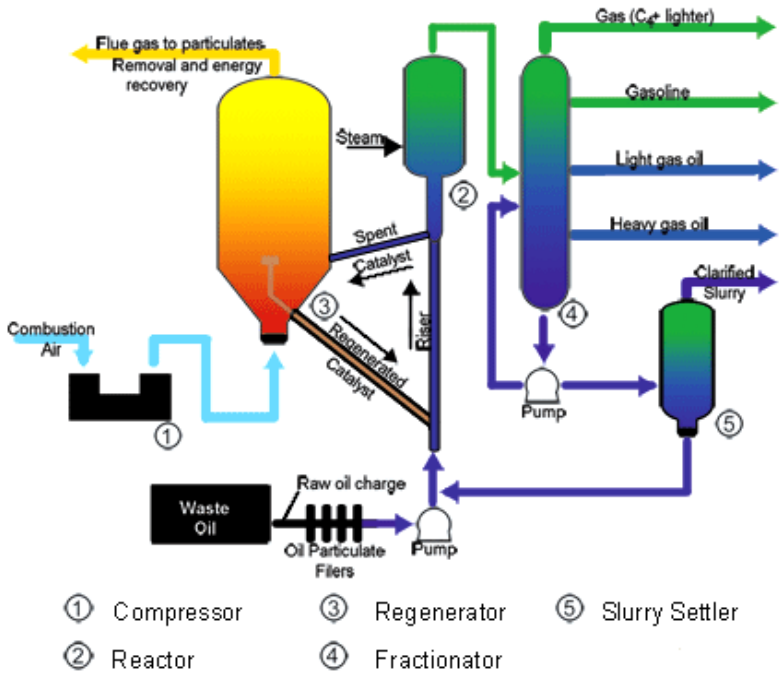
Source: (Tungler, 2012)

Figure 2.4b Flow Diagram of Fluid Catalytic Cracking Process



Source: (Tungler, 2012)

Figure 2.5: FCC Unit Schematic Diagram



Source: (Srikanth, 2011)

Figure 2.6 FCC Unit Scheme Diagram

As can be seen in figure 2.6 modern FCC units are all continuous processes with catalyst regeneration and recycling system inbuilt which operate 24 hours a day for as long as 2 to 3 years between scheduled shutdowns for routine maintenance.

Fluid catalytic cracking (FCC) continues to play a key role in an integrated refinery as the primary conversion process. For many refiners, the catalytic cracker is the key to profitability in that the successful operation of the unit determines whether or not the refiner can remain competitive in today's market. Approximately 350 to 400 cat crackers are operating worldwide, with a total processing capacity of over 12.7 million barrels per day (Hug, 1998). Most of the existing FCC units have been designed or modified by six major technology licensers:

1. ABB Lummus Global

2. Exxon Research and Engineering (ER&E) (Figure 2.7a)

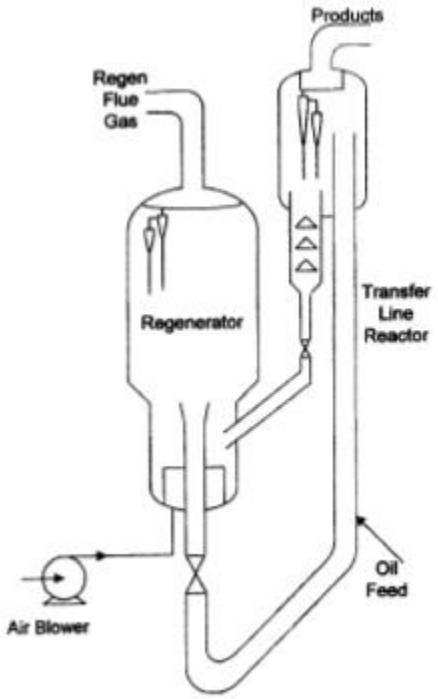
3. Kellogg Brown & Root—KBR (formerly The M.W. Kellogg

Company) (Figure 2.7b)

4. Shell Oil Company

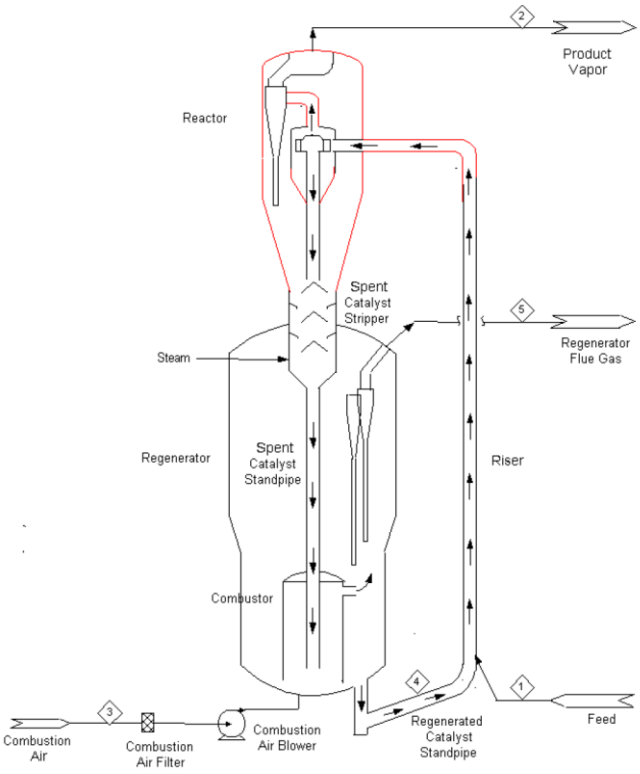
5. Stone & Webster Engineering Corporation (SWEC)/IFP (Figure 2.9)

6. UOP (Universal Oil Products) (Figure 2.8)



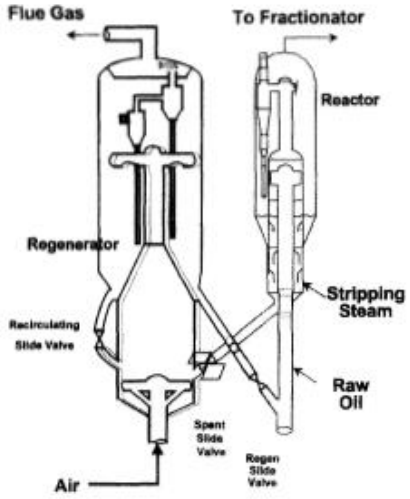
Source: (Bhattacharyya, 2013)

Figure 2.7a Typical schematic of Exxon flexi-cracker



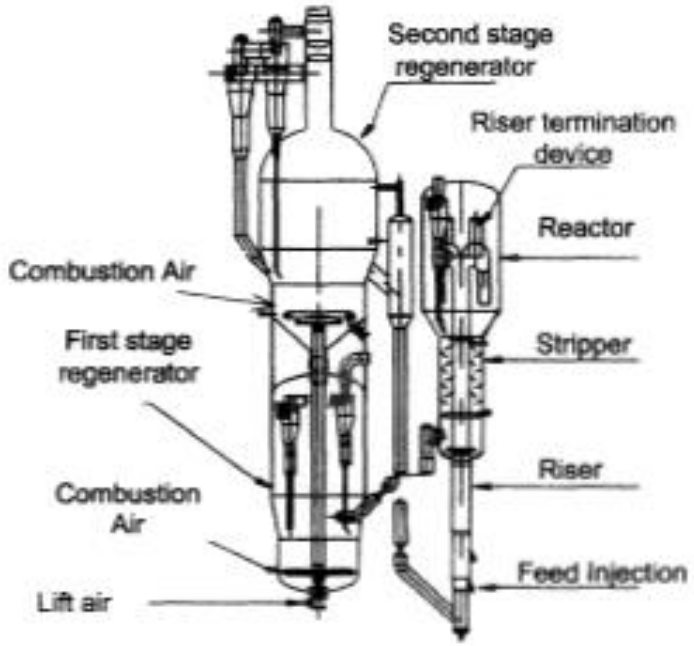
Source: (Ibsen, 2006)

Figure 2.7b Typical schematic of Kellogg Brown & Root—KBR



Source: (Bhattacharyya, 2013)

Figure 2.8 Universal Oil Products (UOP) Fluid Catalytic Cracking Unit



Source: (Bhattacharyya, 2013)

Figure 2.9 SWEC side by side FCC Unit

Figures 2.7a to 2.9 contain sketches of typical unit configurations of some licensers. Although the mechanical configuration of individual FCC units may differ, their common objective is to upgrade low-value feedstock to more valuable products. Worldwide, about 45% of all gasoline comes from FCC and ancillary units, such as the alkylation unit. Since the start-up of the first commercial FCC unit in 1942, many improvements have been made. These improvements have enhanced the unit's mechanical reliability and its ability to crack heavier, lower value feedstock. The FCC has a remarkable history of adapting to continual changes in market demands. The main purpose of the unit is to convert high-boiling petroleum fractions called gas oil to high value, high-octane gasoline and heating oil. Gas oil is the portion of crude oil that commonly boils in the 650+°F to 1,050+°F (330° to 550°C) range.

The gas oil feed for the conventional cat cracker comes primarily from the atmospheric column, the vacuum tower, and the delayed coker. In addition, a number of refiners blend some atmospheric or vacuum residue into the feedstock to be processed in the FCC unit. The FCC process is very complex. For clarity, the process description has been broken down into separate sections:

i- Feed Preheat

ii- Riser—Reactor—Stripper

iii- Regenerator—Heat/Catalyst Recovery

iv- Main Fractionator

2.6.3 FCC converter

The riser cracking converter consists of a riser reactor, disengager, stripper, regenerator, and catalyst circulation facilities (Plate I and figures 2.10a&b) Auxiliary equipment in the section includes;

i- Fresh and used catalyst storage drums.

ii- Air blower

iii- Flue gas orifice chamber

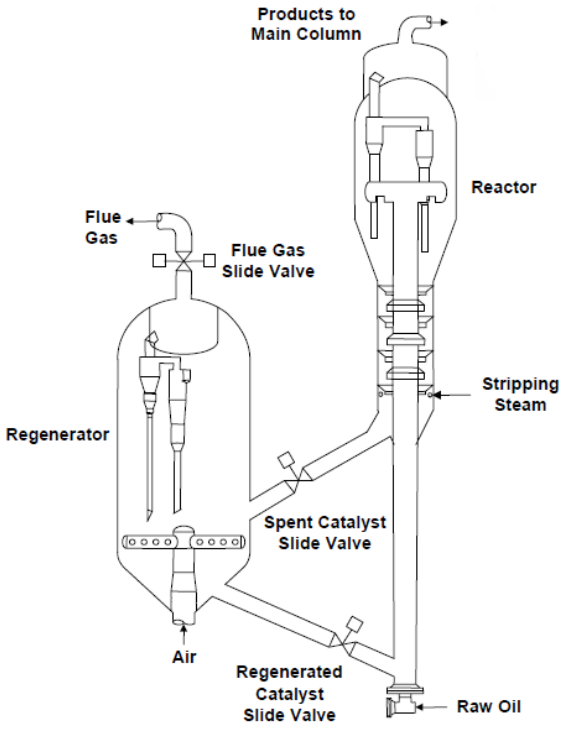
iv- Water seal pot and

v- CO Boiler

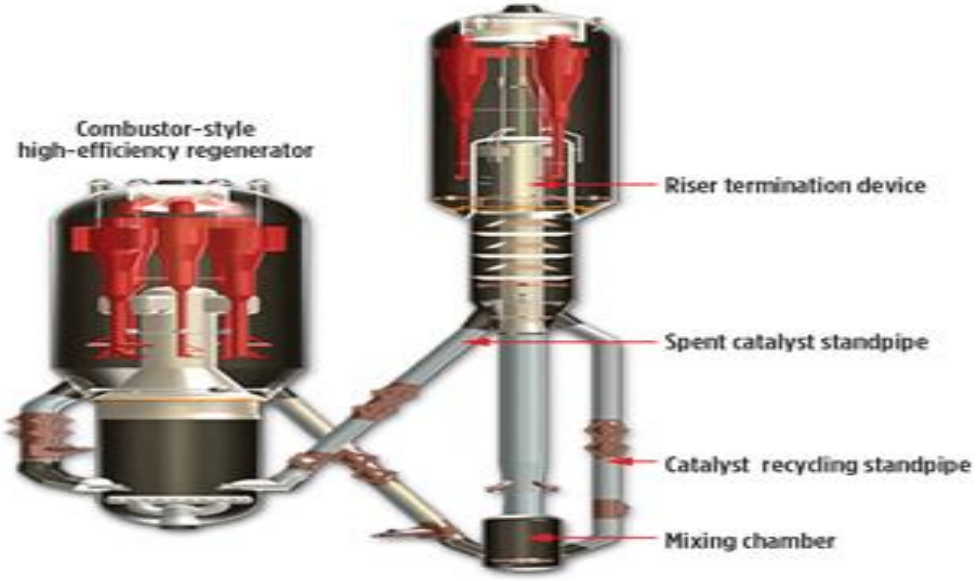


Source: (KRPC FCCU)

Plate I: Pictorial View of Fluid Catalytic Cracking Converter



Source: (Srikanth, 2011)
Figure 2.10a the Converter Schematic



Source: (Bhattacharyya, 2013)

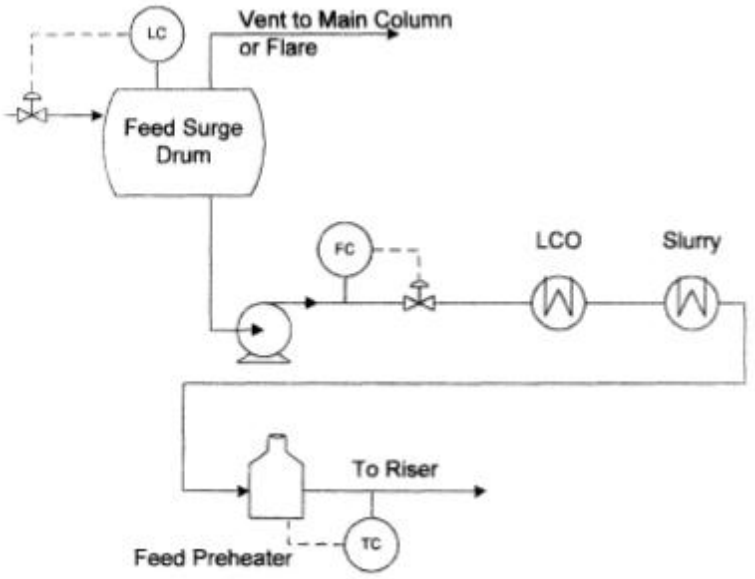
Figure 2.10b the Converter Internals

The Converter is the heart of FCC process. In a modern Catalytic Cracker, Virtually all the reactions occur in 2.1 to 3.0 seconds before the catalyst and the products are separated in the reactor. From the preheater, the feed enters the riser near the base where it contacts the regenerated catalyst; the ratio of catalyst-to-oil is normally in the range of 4:1 to 9:1 by weight. The heat absorbed by the catalyst in the regenerator provides the energy to heat the feed to its desired reactor temperature. The heat of the reaction occurring in the Riser is endothermic. The Circulating catalyst provides this energy. The typical regenerated catalyst temperature ranges between 677 ^{o}C to 732 ^{o}C (KRPC Handbook, 1980).

2.6.4 Feed preheat

Most refineries produce sufficient gas oil to meet the catalytic crackers' demand. However, in those refineries in which the gas oil produced does not meet the catalytic cracker capacity, it may be economical to supplement feed by purchasing FCC feedstocks or blending some residue. The refinery-produced gas oil and any supplemental FCC feedstock are generally combined and sent to a surge drum, which provides a steady flow of feed to the charge pumps. This drum can also separate any water or vapor that may be in the feedstock. From the surge drum, the feed is normally heated to a temperature of 500°F to 700°F (260°C to 370°C). The main Fractionator bottoms pump around and/or fired heaters are the usual sources of heat. The feed is first routed through heat exchangers using hot streams from the main Fractionator. The main Fractionator top pump around, light cycle oil product, and bottoms pump around are commonly used (Figure 2.11). Removing heat from the main Fractionator is at least as important as preheating the feed. Most FCC units use fired heaters for FCC feed final preheat. The feed Preheater provides control over the catalyst-to-oil ratio, a key variable in the process (KRPC Handbook, 1980).

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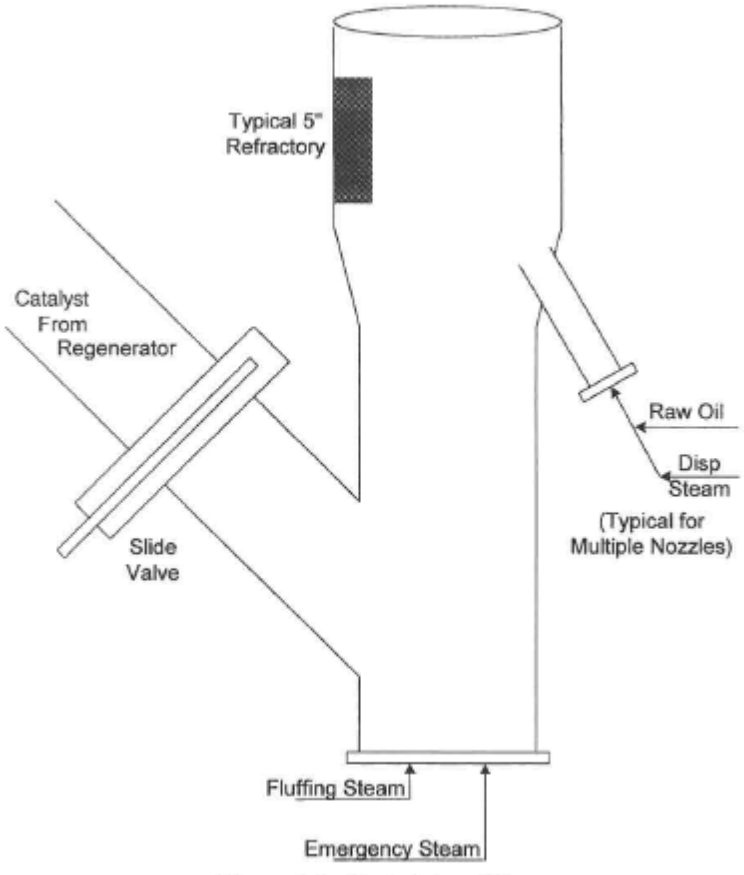


Source: (Reza, 2000)

Figure 2.11 Typical Feed Preheat System

2.6.5 Riser—reactor—stripper

The reactor-regenerator (the converter) is the heart of the FCC process. In a modern catalytic cracker, virtually all the reactions occur in 2.1 to 3.0 seconds before the catalyst and the products are separated in the reactor (Reza, 2000). From the Preheater, the feed enters the riser near the base where it contacts the regenerated catalyst (see Figure 2.12). The ratio of catalyst to- oil is normally in the range of 4:1 to 9:1 by weight (Letzsch et al, 2009). The heat absorbed by the catalyst in the regenerator provides the energy to heat the feed to its desired reactor temperature. The heat of the reaction occurring in the riser is endothermic (i.e., it requires energy input). The circulating catalyst provides this energy. The typical regenerated catalyst temperature ranges between 1,250°F to 1,350°F (677°C to 732°C).



Source: (Reza, 2000)

Figure 2.12 Typical Riser Y

The catalytic reactions occur in the vapor phase. Cracking reactions begin as soon as the feed is vaporized. The expanding volumes of the vapors that are generated are the main driving force to carry the catalyst up the riser. Catalyst and products are quickly separated in the reactor. However, some thermal and non-selective catalytic reactions continue. A number of refineries are modifying the riser termination devices to minimize these reactions. The riser is a vertical pipe. It usually has 4- to 5-inch (10 to 13 cm) thick refractory lining for insulation and abrasion resistance. Typical risers are 2 to 6 feet (60 to 180 cm) in diameter and 75 to 120 feet (25 to 30

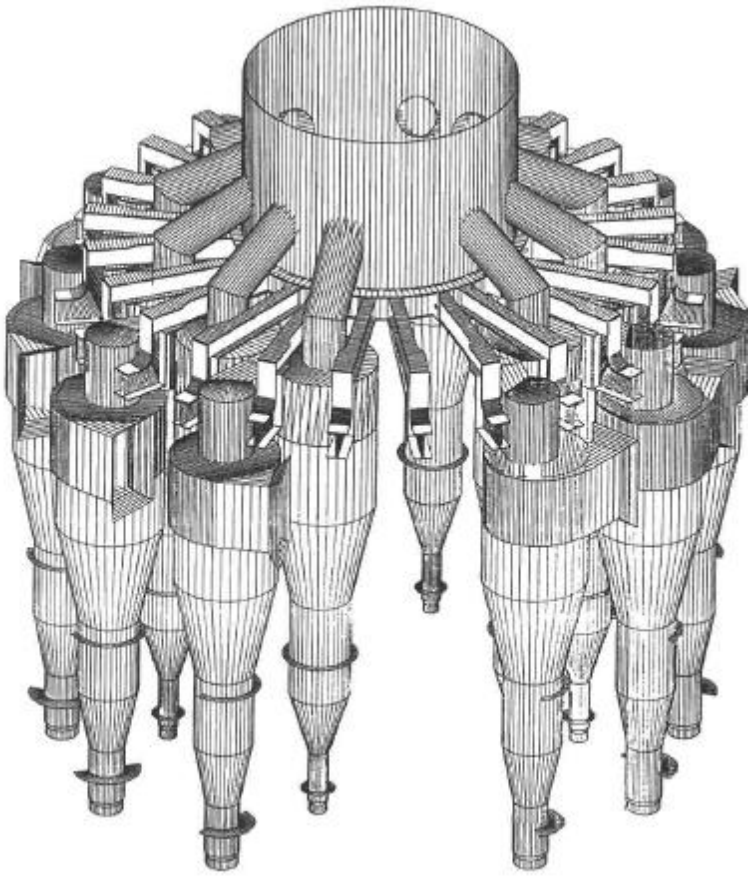
meters) long. The ideal riser simulates a plug flow reactor, where the catalyst and the vapor travel the length of the riser with minimum back mixing (Reza, 2000).

Efficient contacting of the feed and catalyst is critical for achieving the desired cracking reactions. Steam is commonly used to atomize the feed. Smaller oil droplets increase the availability of feed at the reactive acid sites on the catalyst. With high-activity zeolite catalyst, virtually all of the cracking reactions take place in three seconds or less. Risers are normally designed for an outlet vapor velocity of 50 ft/sec to 75 ft/sec (15.2 to 22.8 m/sec). The average hydrocarbon residence time is about two seconds (based on outlet conditions). As a consequence of the cracking reactions, a hydrogen-deficient material called coke is deposited on the catalyst reducing catalyst activity.

2.6.6 Riser catalyst separation

After exiting the riser, catalyst enters the reactor vessel. In today's FCC operations, the reactor serves as housing for the cyclones. In the early application of FCC, the reactor vessel provided further bed cracking, as well as being a device used for additional catalyst separation. Nearly every FCC unit employs some type of inertial separation device connected on the end of the riser to separate the bulk of the catalyst from the vapors (Letzsch et al., 2009). A number of units use a deflector device to turn the catalyst direction downward. On some units, the riser is directly attached to a set of cyclones (Figure 2.13). The term "rough cut" cyclones generally refers to this type of arrangement. These schemes separate approximately 75% to 99% of the catalyst from product vapors. Most FCC units employ either single or two-stage cyclones (Figure 2.13) to separate the remaining catalyst particles from the cracked vapors. The cyclones collect and return the catalyst to the stripper through the diplegs and flapper/trickle valves. The product vapors exit

the cyclones and flow to the main fractionators for recovery. The efficiency of a typical two-stage cyclone system is 99.995+ %



Source: (Reza, 2000)

Figure 2.13 Two stage cyclone system,

It is important to separate catalyst and vapors as soon as they enter the reactor. Otherwise, the extended contact time of the vapors with the catalyst in the reactor housing will allow for non selective catalytic re-cracking of some of the desirable products (Reza, 2000).

2.6.7 Disengager

The cracked vapors, inert, dispersion steam and steam from the spent catalyst stripper flow upward through the disengager and leave through sets of one stage cyclones. These cyclones separate all but small quantities of the entrained catalyst from the vapour and return to the bottom of the disengager. The vapour then leaves the disengager to the Fractionator. The catalyst separated in these cyclones passes through the dip leg to join the catalyst from the rough-cut cyclone.

2.6.8 Stripping section

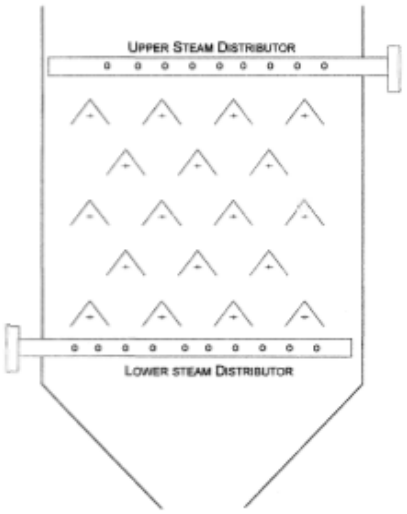
As the spent catalyst falls into the stripper (Figure 2.14a/b), hydrocarbons are adsorbed on the catalyst surface, hydrocarbon vapors fill the catalyst pores, and the vapors entrained with the catalyst also fall into the stripper. Stripping steam, at a rate of 2 to 5 1bs per 1,000 lbs (2 kg to 5 kg per 1,000 kg,) is primarily used to remove the entrained hydrocarbons between catalyst particles. Stripping steam does not address hydrocarbon desorption and hydrocarbons' filling the catalyst pores. However, reactions are driven by the reactor temperature and the catalyst residence time in the stripper. The higher temperature and longer residence time allow conversion of adsorbed hydrocarbons into "clean lighter" products. Both baffled and unbaffled stripper designs (Figure 2.14a/b) are in commercial use. An efficient stripper design generates intimate contact between the catalyst and steam. Reactor strippers are commonly designed for a steam superficial velocity of 0.75 ft/sec (0.23 m/sec) and a catalyst flux rate of 500 to 700 lbs per minute per square foot (2.4 kg to 3.4 kg per minute per square meter). At too high a flux, the falling catalyst tends to entrain steam, thus reducing the effectiveness of stripping steam. It is important to minimize the amount of hydrocarbon vapors carried through to the regenerator, but not all the hydrocarbon vapors can be displaced from the catalyst pores in the stripper. A fraction of them are carried with the spent catalyst into the regenerator. This hydrocarbon vapors/liquid

has a higher hydrogen-to-carbon ratio than the coke on the catalyst. The drawbacks of allowing these hydrogen rich hydrocarbons to enter the regenerator are as follows:

i- Loss of liquid product: Instead of the hydrocarbons burning in the regenerator, they could be recovered as liquid products.

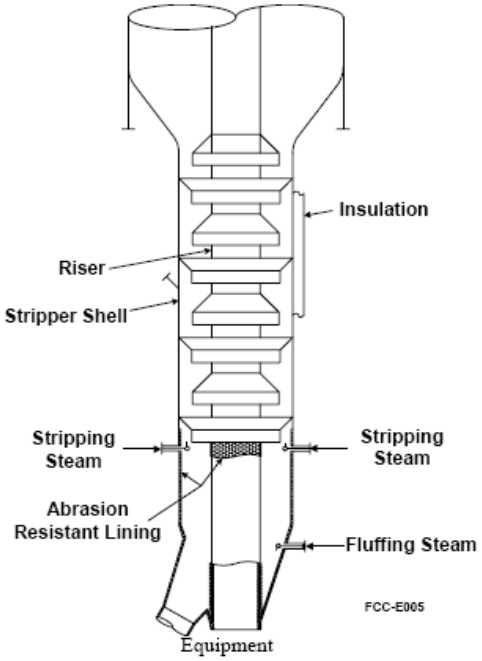
ii- Loss of throughput: The combustion of hydrogen to water produces 3.7 times more heat than the combustion of carbon to carbon dioxide. The increase in the regenerator temperature caused by excess hydrocarbons could exceed the temperature limit of the regenerator internals and force the unit to a reduced feed rate mode of operation.

iii- Loss of catalyst activity: The higher regenerator temperature combined with the formation of steam in the regenerator reduces catalyst activity by destroying the catalyst's crystalline structure. The flow of spent catalyst to the regenerator is typically controlled by a valve that slides back and forth. This slide valve is controlled by the catalyst level in the stripper. The catalyst height in the stripper provides the pressure head, which allows the catalyst to flow into the regenerator. The exposed surface of the slide valve is usually lined with refractory to withstand erosion (Figure 2.14). In a number of earlier FCC designs, lift air is used to transport the spent catalyst into the regenerator



Source: (Reza, 2000)

Figure 2.14a Example of a two stage stripper

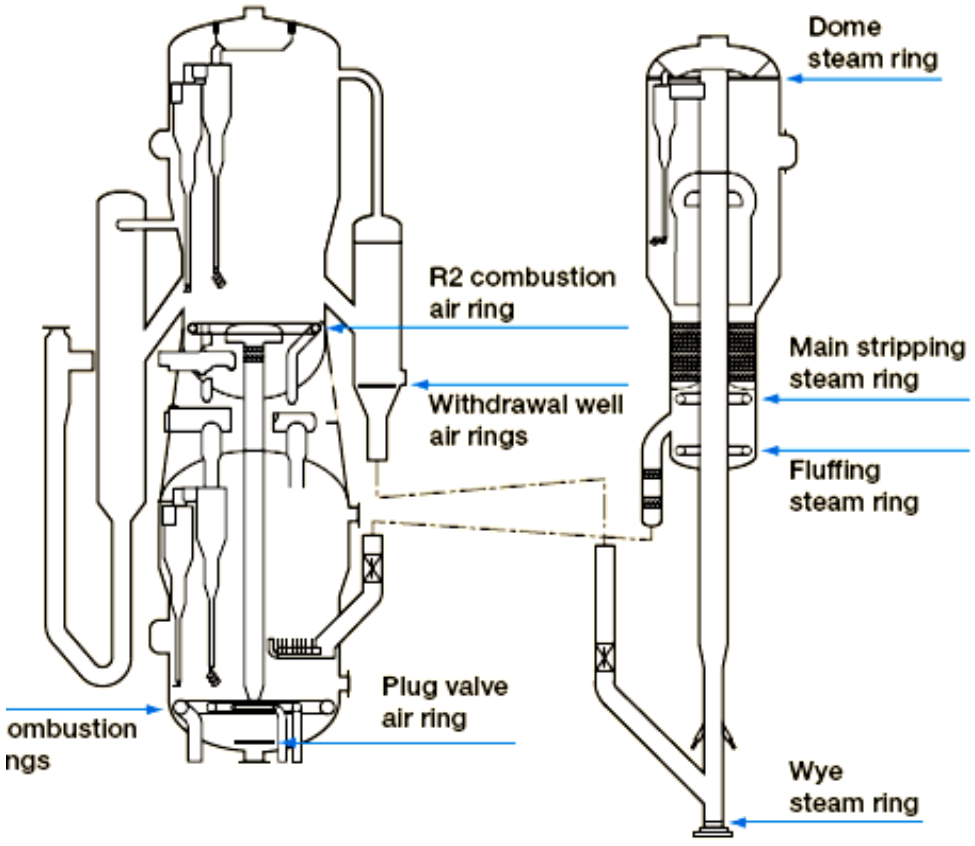


Source: (Srikanth, 2011)

Figure 2.14b Catalyst Stripper

2.6.9 Steam ring applications

The steam ring distributor has several potential applications in the reactor/stripper section of the FCC unit. The main stripping steam ring, located in the stripper, is designed to displace entrained hydrocarbons from the spent catalyst. One or more of these rings may be necessary, based on the configuration and diameter of the stripper (Figure 2.15). Recent designs often split the ring into two half-rings. This design allows for greater flexibility in the operation of each half-ring, which is helpful in cases where catalyst flows preferentially down one side of the vessel.



Source: (Cooper, 2011)

Figure 2.15 Typical Rings with a Shaw-Designed Residue Catalytic Cracker



Source: (Cooper, 2011)

Plate II: Air Grid in Fabrication Shop

2.7 Catalyst Regeneration

Catalyst can be regenerated over a range of temperatures and flue gas composition with inherent limitations. Two distinctive different modes of regeneration are practiced: Partial Combustion and complete combustion. Complete combustion generates more energy when coke yield is increased; Partial combustion generates less energy when the coke yield is increased. In complete combustion, the excess reaction component is oxygen, so more carbon generates more combustion. In partial combustion, and an increase in coke yield means a shift from CO_2 to CO.

FCC regeneration can be further sub divided into Low, Intermediate and high temperature regeneration. In low temperature regeneration about 640 ^{0}C, complete combustion is impossible. One of the characteristics of low temperature regeneration is that at 1,190 ^{0}F, all three components (O_2, CO and CO_2) are present in the flue gas at significant levels. Low temperature regeneration was the mode of operation that was used in the early implementation of the catalytic cracking process.

High temperature regeneration was developed in early 1970s (Reza, 2000). High temperature regeneration meant increasing the temperature until all the oxygen was burned. The main result

was low carbon on the regenerated catalyst. This mode of regeneration required maintaining in the flue gas, either a small amount of excess oxygen or no CO, or in excess oxygen and a variable quantity of CO. If there was excess oxygen, the operation was in partial burn.

With the advent of combustion promoter, the regenerated temperature could be reduced and still maintain full burn. Thus intermediate temperature regeneration is not necessary stable unless combustion promoter is used to assist in the combustion of CO in the dense phase Table 2.4 contains a 2x3 matrix summarizing various aspect of regeneration.

The following matrix of regeneration temperatures and operating modes shows the inherent limitations of operating regions, regeneration is either partial or complete, at low intermediate or high temperatures.

Table 2.4 Regenerator Temperatures and Operating Modes

Operating Regions of Regenerator Combustion	**Partial Combustion Mode**	**Full Combustion Mode**
Low temperature (Normally 640 ^{0}C)	Stable (Small Afterburning) O_2,CO_2 and CO in the flue Gas	Not achievable
Intermediate temperature (normally 690 ^{0}C)	Stable (with combustion promoter); tends to have high carbon on regenerated catalyst	Stable with combustion Promoter
High temperature (normally 730 ^{0}C)	Stable operation	Stable operation

Source: (Reza, 2000).

At low temperatures, regeneration is always partial, carbon on regenerated catalyst is high, and increasing combustion air results in after burn. At intermediate temperatures, carbon on regenerated catalyst is reduced. The three normal Operating Regions are indicated in table 2.4 above.

There are some advantages and disadvantages associated with full and partial combustion:

2.7.1 Advantages of complete combustion

i- Energy Efficient

ii- Heat balances at low coke yield

iii- Minimum Hardware no CO boiler

iv- Better yield from clean feed

2.7.2 Disadvantages of complete combustion

i- Narrow range of coke yields unless some heat removal system is incorporated

ii- Greater afterburn, particularly with an uneven air or spent catalyst distribution system

iii- Low Catalyst/oil Ratio

iv- The Choice of partial versus full combustion is dictated by FCC feed quality, with "clean Feed", Full combustion is the choice. With low quality feed or resid, partial combustion, possibly with heat removal, is the choice (Rader, 1996).

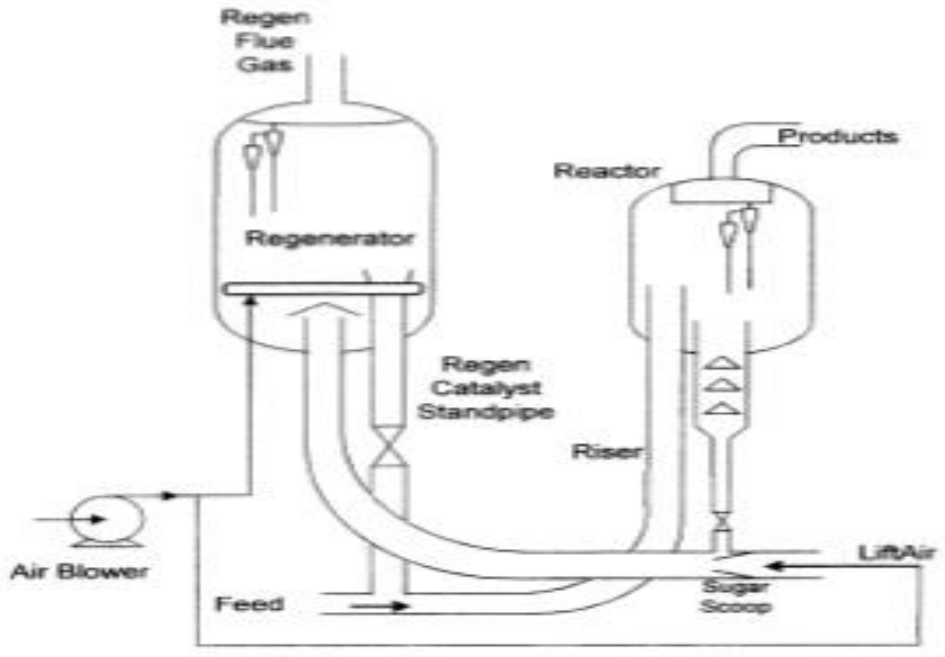
2.7.3 Regenerator–heat/catalyst recovery

The regenerator (Figure 2.16) has two main functions: it restores catalyst activity and supplies heat to crack the feed by using hot air to burn off the coke with the aid of an air grid (plate II). The spent catalyst entering the regenerator contains between 0.4 wt% and 2.5 wt% coke, depending on the quality of the feedstock. Components of coke are carbon, hydrogen, and a trace of sulfur and nitrogen. This burns according to the following reactions shown in Table 2.5.

Table 2.5 Reactions occurring in the regenerator

REACTIONS			K Cal/kg of C, H_2, or S	BTU/Ib of C, H_2, or S
C + ½ O_2	⟶	CO_2	2,200	3,968
CO + ½ O_2	⟶	CO_2	5,600	10,100
C + O_2	⟶	CO_2	7,800	14,100
H2 + ½ O_2	⟶	H_20	28,900	52,125
S + XO	⟶	SO_X	2,209	3,983
N + XO	⟶	NO_X	-	-

Source:(Reza 2000).



(Reza, 20000)

Figure 2.16: A typical Regenerator using Lift air to transfer Catalyst

Air provides oxygen for the combustion of coke and is supplied by one or more air blowers. The air blower provides sufficient air velocity and pressure to maintain the catalyst bed in a fluid state. The air enters the regenerator through an air distributor (Figure 2.15) located near the bottom of the vessel. The design of an air distributor is important in achieving efficient and reliable catalyst regeneration (plate II). Air distributors are typically designed for a 1.0 psi to 2.0 psi (7 to 15 Kpa) pressure drop to ensure positive air flow through all nozzles. There are two regions in the regenerator: the dense phase and the dilute phase. At velocities common in the regenerator, 2 ft/sec to 4 ft/sec 0.6 to 1.2 m/sec), the bulk of catalyst particles are in the dense bed immediately above the air distributor. The dilute phase is the region above the dense phase up to the cyclone inlet, and has a substantially Lower catalyst Concentration (Reza, 2000).

2.7.4 Regenerator bed temperature

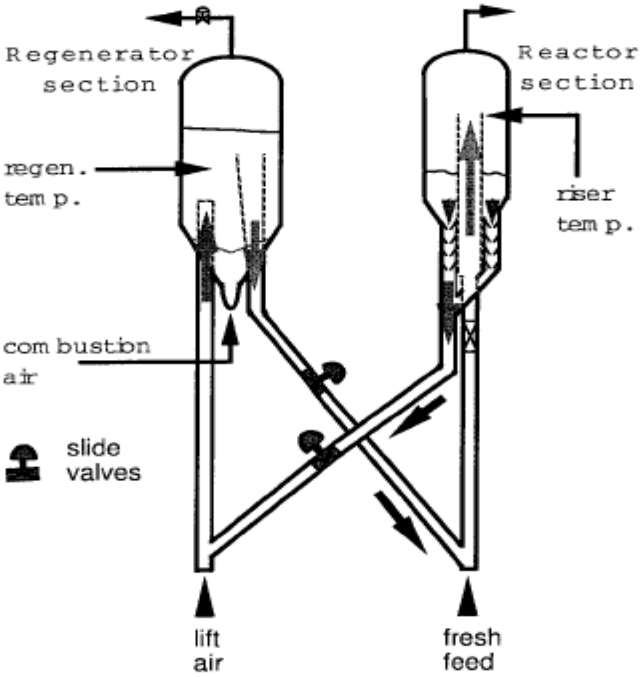
Regenerator dense bed temperature varies slightly along the height of the bed (Reza, 2000) being lowest at the bottom and highest at the top. If generator bed temperature is too low there is insufficient combustion of coke leading to coke buildup on catalyst which will reduce overall cracking efficiency. On the other hand, excessive regenerator temperature will result in rapid thermal deactivation of the catalyst (Wilczura-Wachnik, 1973).

2.7.5 Standpipe/slide valve

During regeneration, the coke level on the catalyst is typically reduced to 0.05%. From the regenerator, the catalyst flows down a transfer line commonly referred to as a standpipe. The standpipe provides the necessary pressure to circulate the catalyst around the unit. Some standpipes extend into the regenerator, and the top section is often called a catalyst hopper. The

hopper, internal to the regenerator, is usually an inverted cone design. In units with "long" catalyst standpipes, external withdrawal hoppers are often used to feed the standpipes. The hopper provides sufficient time for the regenerated catalyst to be "de-bubbled" before entering the standpipe.

Standpipes are typically sized for a flux rate in the range of 100 to 300 lb/sec/ft^2 (500 to 1,500 kg/sec/m^2) of circulating catalyst. In most cases, sufficient flue gas is carried down with the regenerated catalyst to keep it fluidized. However, longer standpipes may require external aeration to ensure that the catalyst remains fluidized. A gas medium, such as air, steam, nitrogen, or fuel gas, is injected along the length of the standpipe. The catalyst density in a well-designed standpipe is in the range of 35 – 45lb/ft^3 (560 - 720kg/m^3). The flow rate of the regenerated catalyst to the riser is commonly regulated by either a slide or plug valve. The operation of a slide valve is similar to that of a variable orifice. Slide valve operation is often controlled by the reactor temperature. Its main function is to supply enough catalyst to heat the feed and achieve the desired reactor temperature (KRPC Handbook, 1980). In Exxon Model IV and flexi-cracker designs, the regenerated catalyst flow is mainly controlled by adjusting the pressure differential between the reactor and regenerator (Ye-Mon, 2004)



Source: (Hug, 1998)

Figure 2.17 FCCU Schematic Slide Valve Installed

2.7.6 Regenerator catalyst separation

As flue gas leaves the dense phase of the regenerator, it entrains catalyst particles. The amount of entrainment largely depends on the flue gas superficial velocity. The larger catalyst particles fall back into the dense bed. The smaller particles are suspended in the dilute phase and carried into the cyclones. Most FCC unit regenerators employ 4 to 16 parallel sets of primary and secondary cyclones. The cyclones are designed to recover catalyst particles greater than 20 microns diameter (Letzsch et al., 2009). The recovered catalyst particles are returned to the regenerator via the diplegs. The distance above the catalyst bed at which the flue gas velocity has stabilized is referred to as the transport disengaging height (TDH). At this height, the catalyst concentration in the flue gas stays constant; none will fall back into the bed. The centerline of the first-stage

cyclone inlets should be at TDH or higher; otherwise, excessive catalyst entrainment will cause extreme catalyst losses.

2.7.7 Flue gas heat recovery schemes

The flue gas exits the cyclones to a plenum chamber in the top of the regenerator. The hot flue gas holds an appreciable amount of energy. Various heat recovery schemes are used to recover this energy. In some units, the flue gas is sent to a CO boiler (Plate III) where both the sensible and combustible heat is used to generate high-pressure steam. In other units, the flue gas is exchanged with boiler feed water to produce steam via the use of a shell/tube or box heat exchanger. In most units, about two-thirds of the flue gas pressure is let down via an orifice chamber or across an orifice chamber. The orifice chamber is vessel containing a series of perforated plates designed to maintain a given backpressure upstream of the regenerator pressure control valve. In some larger units, a turbo expander is used to recover this pressure energy. To protect the expander blades from being eroded by catalyst, flue gas is first sent to a third-stage separator to remove the fines. The third-stage separator, which is external to the regenerator, contains a large number of swirl tubes designed to separate 70% to 95% of the incoming particles from the flue gas.



Source: (KRPC FCCU)

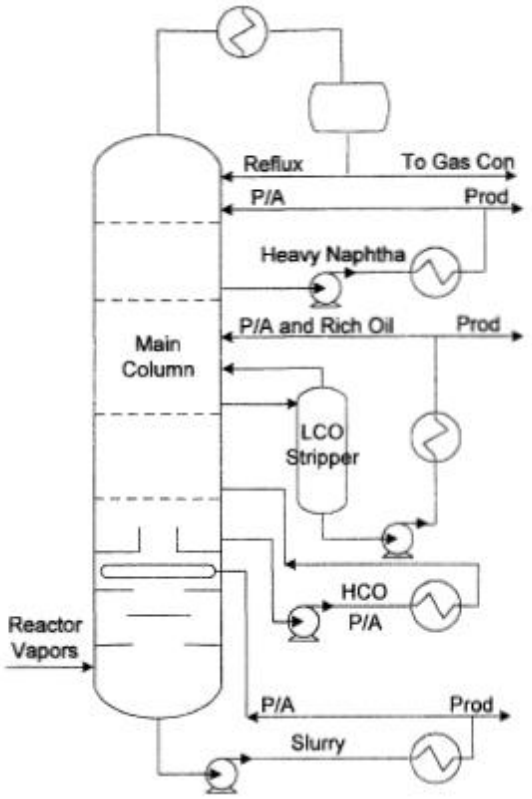
Plate III: Pictorial View of CO Boiler Unit

A power recovery train employing a turbo expander usually consists of four parts: the expander, a motor/generator, an air blower, and a steam turbine. The steam turbine is primarily used for start-up and, often, to supplement the expander to generate electricity. The motor/generator works as a speed controller and flywheel; it can produce or consume power. In some FCC units,

the expander horsepower exceeds the power needed to drive the air blower and the excess power is output to the refinery electrical system. If the expander generates less power than what is required by the blower, the motor/generator provides the power to hold the power train at the desired speed. From the expander, the flue gas goes through a steam generator to recover thermal energy. Depending on local environmental regulations, an electrostatic precipitator (ESP) or a wet gas scrubber may be placed downstream of the waste heat generator prior to release of the flue gas to the atmosphere.

2.8 Fractionator

The purpose of the main Fractionator, or main column (Figure 2.18), is to de-superheat and recover liquid products from the reactor vapors. The hot product vapors from the reactor flow into the main fractionators near the base. Fractionation is accomplished by condensing and re-vaporizing hydrocarbon components as the vapor flows upward through trays in the tower. The operation of the main column is similar to a crude tower, but with two differences. First, the reactor effluent vapors must be cooled before any fractionation begins. Second, large quantities of gases will travel overhead with the Unstabilized gasoline for further separation. The bottom section of the main column provides a heat transfer zone. Shed decks, disk/doughnut trays, and grid packing are among some of the contacting devices used to promote vapor/liquid contact. The overhead reactor vapor is de-superheated and cooled by a pump around stream. The cooled pump around also serves as a scrubbing medium to wash down catalyst fines entrained in the vapors. Pool quench can be used to maintain the Fractionator bottoms temperature below coking temperature, usually at about 700°F (370°C). The recovered heat from the main column bottoms is commonly used to preheat the fresh feed, generate steam, and serve as a heating medium for the Gas plant reboilers.



Source: (Reza, 2000).

Figure 2.18 FCC typical Fractionator Circuit

The heaviest bottoms product from the main column is commonly called slurry or decant oil, the decant oil is often used as a "cutter stock" with vacuum bottoms to make No. 6 fuel oil. High-quality decants oil (low sulfur, low metals, and low ash) can be used for carbon black feedstocks. Early FCC units had soft catalyst and inefficient cyclones with substantial carryover of catalyst to the main column where it was absorbed in the bottoms. Those FCC units controlled catalyst losses in two ways. First, they used high recycle rates to return slurry to the reactor. Second, the slurry product was routed through slurry settlers, either gravity or centrifugal, to remove catalyst fines. A slipstream of FCC feed was used as a carrier to return the collected fines from the separator to the riser. Since then, improvements in the physical properties of FCC catalyst and in

the reactor cyclones have lowered catalyst carry-over. Most units today operate without separators. The decant oil is sent directly to the storage tank. Catalyst fines accumulate in the tank, which is cleaned periodically. Some units continue to use some form of slurry settler to minimize the ash content of decanted oil. Above the bottoms product, the main column is often designed for three possible side cuts:

i- Heavy cycle oil (HCO)—used as a pump around stream, sometimes as recycle to the riser, but rarely as a product
ii- Light cycle oil (LCO)—used as a pump around stream, sometimes as absorption oil in the gas plant, and stripped as a product for diesel blending; and
iii- Heavy naphtha—used as a pump around stream, sometimes as absorption oil in the gas plant, and possible blending in the gasoline pool.

In many units, the light cycle oil (LCO) is the only side cut that leaves the unit as a product. LCO is withdrawn from the main column and routed to a side stripper for flash control. LCO is sometimes treated for sulfur removal prior to being blended into the heating oil pool. In some units, a slipstream of LCO, either stripped or unstripped, sent to the sponge oil absorber in the gas plant. In other units, sponge oil is the cooled, unstripped LCO. Heavy cycle oil, heavy naphtha, and other circulating side pump around reflux streams are used to remove heat from the Fractionator. They supply reboil heat to the gas plant and generate steam. The amount of heat removed at any pump around point is set to distribute vapor and liquid loads evenly throughout the column and to provide the necessary internal reflux. Unstabilized gasoline and light gases pass up through the main column and leave as vapor. The overhead vapor is cooled and partially condensed in the Fractionator overhead condensers. The stream flows to an overhead receiver, typically operating at <15 psig (<1 bar) (KRPC Handbook, 1980).

Hydrocarbon vapor, hydrocarbon liquid, and water are separated in the Drums. Hydrocarbon vapors flow to the wet gas compressor. This gas stream contains not only ethane and lighter gases, but about 95% of the C_3 and C_4 and about 10% of the naphtha. The phrase "wet gas" refers to condensable components of the gas stream.

2.8.1 Products and equipments involved in fractionation

2.8 .1.1 Bottom product

Part of the product is recycled to the converter to recover catalyst dusts. The slurry is also recycled to the converter according to conversion requirement. The remaining bottom product passes into the stripper to eliminate the light fractions and then, after cooling, is sent to storage.

2.8.1.2 Light cycle oil

The light cycle oil is collected in a draw off tray of the main Fractionator, sent to the stripping column and then sent to storage.

2.8.1.3 Gas – light naphtha – heavy naphtha

After cooling and condensation by air coolers, the top vapours are collected in a separator. The condensed water is sent from the separator to the FCC sour water stripper section. Part of the liquid hydrocarbons is refluxed at the top of the column and the rest, through a water cooler, goes into a second separator together with the gaseous phase. The liquid hydrocarbons are pumped from this separator into the absorption column. The incondensable products are collected by a compressor and sent to the absorption column, where the heavy naphtha coming from the Fractionator also goes.

2.8.1.4 The absorber

The absorber is used to maximize recovery of the C_3 fraction from the gas phase. This recovery occurs when the gas phase is forced, under conditions of high pressure and low temperature, through relatively heavy hydrocarbons.

The absorption column has 40 trays and an intermediate reflux. The gas phase enters the lower zone of the column and leaves from the top with a minimum content of C_3 produced from Dealkylation reaction (Table 2.6), it is then sent to gas treating unit.

The light naphtha, coming from the main tower overhead separator, enters at about halfway up the column while the heavy naphtha enters the top zone. The naphtha, containing the absorbed gases, is sent from the bottom of the column through a further separator into the ethane strippers.

2.8.1.5 Ethane stripper

Maximum recovery of C_3 causes an excessive recovery of C_2 which means that the LPG would not meet specifications. The stripper eliminates the excess of C_2 from the product (naphtha +LPG) to be fed the debutanizer column. The stripper is a column with 20 trays and a bottom re-boiler. The product to be stripped enters at the top, flows down through the hot zone in counter-current to the vapour and, from the bottom, passes into the debutanizer. The C_2 instead goes back into the absorption column.

2.8.1.6 Debutanizer

The debutanizer is a naphtha stabilization column with recovery of C_3 and C_4 fractions. The column has 30 trays and is equipped with a bottom re-boiler and a top reflux. The fractionating system is the normal one. Naphtha after preheating enters the column about half way up. The heat make-up for distillation is supplied by the bottom re-boiler while the top temperature is controlled by reflux. The overhead vapours, condensed in air coolers, are collected in the overhead accumulator and, from here; a part of them is refluxed into the column. After further cooling, the product passes to the gas concentration unit. The bottom product, stabilized gasoline, is cooled and then passes to the MEROX treating unit. All important reactions involve in Fluid Catalytic Cracking are illustrated in Table 2.6.

Table 2.6 Important Reactions occurring in FCC

	Reactions	
1	**Cracking**	
	Paraffins cracked to olefins and smaller Paraffins	$C_{10}H_{22} \rightarrow C_4H_{10} + C_6H_{12}$
	Olefins cracked to smaller olefins Aromatic side chain scission	$C_9H_{18} \rightarrow C_4H_8 + C_5H_{10}$
	Aromatic Side chain-scission Naphthenes (Cycloparaffins) cracked to olefins and smaller ring compounds	$ArC_{10}H_{21} \rightarrow C_4H_8 + C_5H_{10}$ $Cyclo\text{-}C_{10}H_{20} \rightarrow C_6H_{12} + C_4H_8$
2	**Isomerization**	
	Olefin bond shift	$i\text{-}C_4H_8 \rightarrow trans\text{-}2\text{-}C_4H_8$
	Normal Paraffins to iso-olefin	$n\text{-}C_5H_{10} \rightarrow iso\text{-}C_5H_{10}$
	Cyclo-Hexane to Cyclo-pentane	$n\text{-}C_4H_{10} \rightarrow iso\text{-}C_4H_{10}\ C_6H_{12} + C_5H_9CH_3$
3	**Hydrogen Transfer**	Naphthenes + Olefin Aromatic + paraffin
	Cyclo-aromatization	$C_6H_{12} + 3C_5H_{10} \rightarrow C_6H_6 + 3C_5H_{12}$
4	**Trans-alkylation/Alkyl-group Transfer**	$C_6H_4(CH_3)_2 + C_6H_6 \rightarrow 2C_6H_5CH_3$
5	**Cyclization of Olefins to Naphthenes**	$C_7H_{14} \rightarrow CH_3\text{-}cyclo\text{-}C_6H_{11}$
6	**Dehydrogenation**	$n\text{-}C_8H_{18} \rightarrow C_8H_{16} + H_2$
7	**Dealkylation**	$Iso\text{-}C_3H_7\text{-}C_6H_5 \rightarrow C_6H_6 + C_3H_6$
8	**Condensation**	$Ar\text{-}CH = CH_2 + R_1CH = CHR_2 \rightarrow Ar - Ar + 2H$

Kaduna Refining and Petrochemical Company (KRPC, 1980)

2.8.2 Vapor recovery section

The vapor from the compressor suction drum flows to a two-stage centrifugal compressor which is driven by a steam turbine. Two-stage compression is provided to achieve the required discharge pressure with minimum horse power and without excessive temperature. Wash water is injected upstream of the intercoolers after which the vapour and liquid are separated in the interstage receiver. The hydrocarbon liquid is joined with the second stage compressor discharge vapour and this combined steam receives a final injection of water to complete removal of soluble salt by water washing. Stripper overhead vapours and absorber rich oil joins the stream and flow through the high pressure condenser to the high pressure separator. This high pressure flash constitutes a cooled equilibrium stage and improves the recovery efficiency of the absorber-stripper system.

The hydrocarbon liquid from the Fractionator overhead drum passes to the middle section of the absorber tower. This absorber recovers a major part of the C_3's and C_4's in the vapour leaving the high pressure separator. Unstripped heavy naphtha from the Fractionator serves as the lean oil and enters the top tray of the absorber. The rich oil can be returned to the Fractionator but normally will overflow tray 20 of the absorber and pass to the bottom section of this tower(KRPC Handook, 1980).

The off gas from the top of the absorber flows to the fuel gas system. The hydrocarbon liquid in the high pressure separator is pumped to the top tray of the stripper. This tower, re-boiled by the Fractionator heavy naphtha pump around, serves to control the C_2 content of the LPG product and its bottom flow represents the total liquid recovery of the system. This bottom liquid is fed to the debutanizer tower after being heated by exchange with debutanizer bottoms liquid (KRPC Handbook, 1980).

The debutanizer tower is re-boiled by heavy cycle oil pump around from the Fractionator. Reflux and net tower overhead is pumped to battery limit for further treating. The bottoms are cooled first by heat exchange with the feed, and then further cooled by air and water coolers. The product, gasoline, will be treated in a merox unit. Provision is made for recycling debutanizer bottoms to the Fractionator overhead drum to increase the lean oil available to the absorber.

2.9 Nigerian Fluid Catalytic Cracking Units

Nigeria has (3) Refineries that contain catalytic cracking as a unit, namely:

1. Kaduna Refining and Petrochemical Company, (KRPC)
2. Warri Refinery and Petrochemical Company (WRPC)
3. Port Harcourt Refinery Company Limited (PHRC)

Kaduna and Warri Refineries uses Kellogg Brown & Root (KBR) stacked design license while Port Harcourt Refinery uses UOP side by side design license. The installed capacities of the plants and their respective FCC capacities are shown in Table 2.7:

Table 2.7: Nigerian Refineries and their respective FCC Capacities

	KRPC	WRPC	PHRC
Installed Capacity (bpsd)	110,000	125,000	60,000 OLD (150,000 NEW)
FCCU Capacity (m^3/hr)	133	171	250
Licensor	Kellog	Kellog	UOP

Source: (KRPC Handbook, 1980).

The design of Kaduna and Warri Refineries is a two stage Regeneration – Partial Combustion configuration yielding flue gas rich in carbon monoxide CO which is further burnt in the CO Boiler (plate III) to generate High Pressure Steam (HP Steam) from High Pressure Boiler Feed Water-WBH. While the design of Port Harcourt Refinery is a Single Stage Regeneration – Total Combustion configuration yielding Flue Gas rich in Carbon dioxide CO_2. The heat of reaction generated is imparted on an array of tubes - Waste Heat Boiler to generate High Pressure Steam from High Pressure Boiler Feed Water (WBH).

2.10 Reactor design

Chemical reactors are vessels designed to contain chemical reactions. It is the site of conversion of raw materials into products and is also called the heart of a chemical process. The design of a chemical reactor where long chain hydrocarbons would be cracked to finished products on a commercial scale would depend on multiple aspects of chemical engineering. Since it is a very vital step in the overall design of a process, designers ensure that the reaction proceeds with the highest efficiency towards the desired output, producing the highest yield of product in the most cost effective way.

Reactors are designed based on features like mode of operation or types of phases present or the geometry of reactors. They are thus called:

i- Batch or Continuous depending on the mode of operation.

ii- Homogeneous or Heterogeneous depending upon the phases present.

They may also be classified as:

i- Stirred Tank Reactor

ii- Tubular Reactor

iii- Packed Bed Reactor

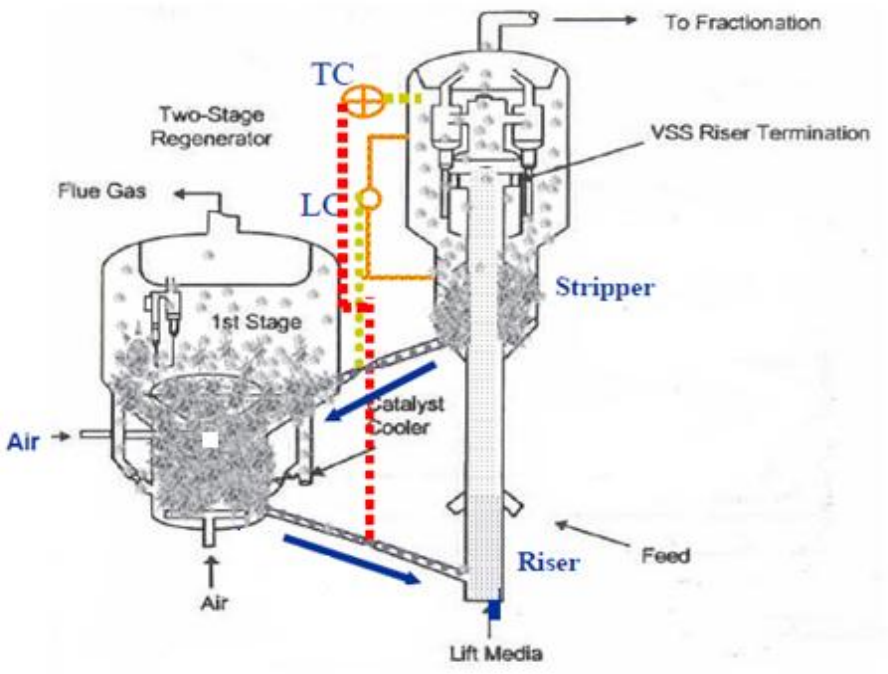
iv- Fluidized Bed Reactor

2.10.1 Fluidized bed reactor

A fluidized bed reactor (FBR) is a type of reactor device that can be used to carry out a variety of multiphase chemical reactions. In this type of reactor, a fluid (gas or liquid) is passed through a granular solid material (usually a catalyst) at high velocities enough to suspend the solid and cause it to behave as though it were a fluid. Uniform particle mixing, uniform temperature gradients and ability to operate reactor in continuous state are some of its advantages which make it one of the widely used reactor in today's industrial world (Sanju, 2010).

2.10.2 Behavior of fluidized bed

Fluidization is an operation by which the beds of solid particles are transformed into a fluid like state through contact with a fluid. The fluidized bed offers the advantage of proper mixing over other contacting methods in which the bed materials are fully suspended in a fluid stream. This is the phenomenon that is established in fluid catalytic reactors as shown in Figure 2.19 below:



Source: (Srikanth, 2011)

Figure 2.19: Fluidization regime in FCC

Figure 2.19 illustrate fluidization regimes in FCC where sections like the Regenerator, Riser and the Disengager are shown. This three sections represent fluidization points in the FCC Converter, as can be seen fluidization is achieved in the riser with the aid of a lift steam while in the disengager separation of vapour from the catalyst with the aid of cyclone and gravity make fluidization possible. In the regenerator hot regeneration air used in burning off coke deposit on the catalyst make fluidization possible thereby separating flue gas from regenerated catalyst. Several factors affect the bed behavior out of which the following will be discussed with regard to the quality of fluidization.

i- Minimum Fluidization Velocity (u_{mf})

ii- Bed Pressure Drop (ΔP)

iii- Bed Expansion Ratio (R)

iv- Bed Fluctuation Ratio (r)

v- Fluidization quality (FQ)

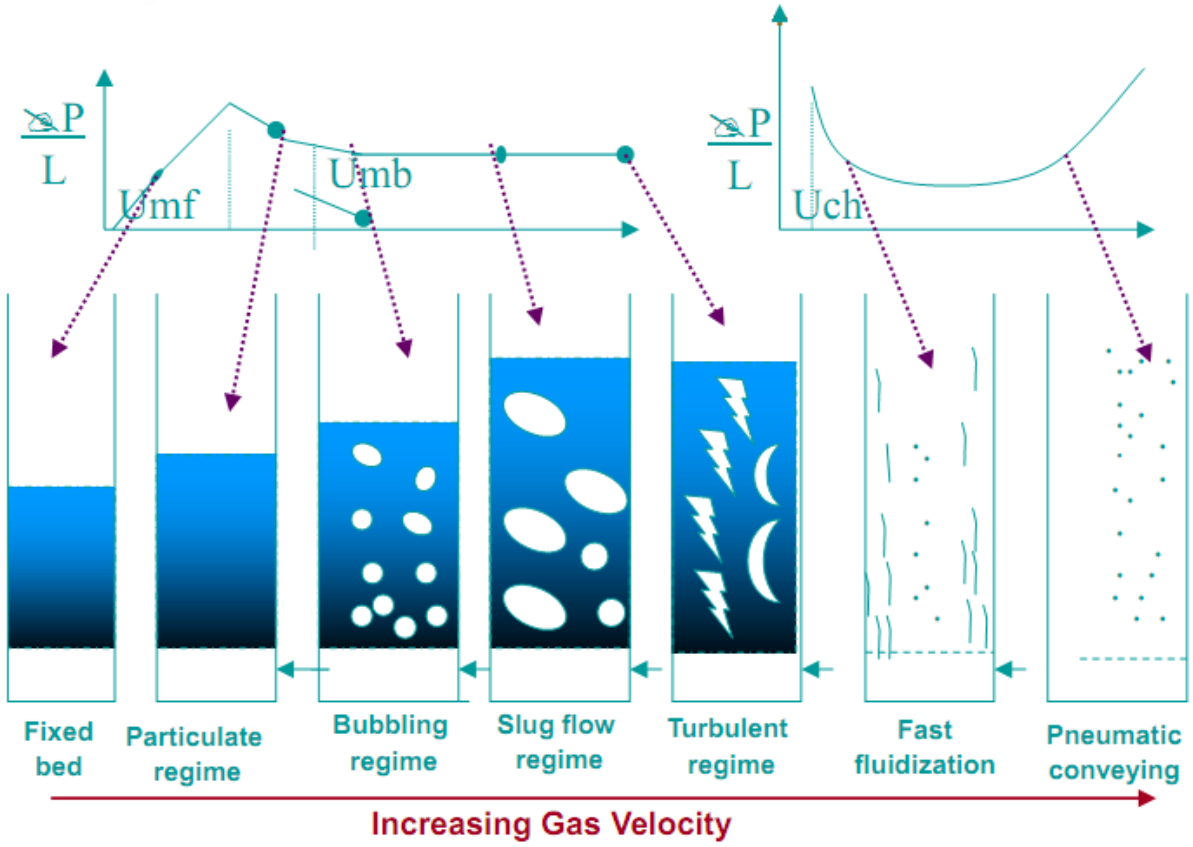
2.10.2.1 Minimum fluidization velocity

When a fluid passes upwards through the interstices of a bed of solids without the slightest disturbance of the solids, the bed is called a fixed bed. With further increase in the velocity of fluid, the entire bed of solids is suspended and behaves as if its weight is counterbalanced by the force of buoyancy (Hartog et al., 2008). At this point, all the particles are fully suspended in the fluid. This is called the onset of fluidization and the velocity of fluid at which it happens is called minimum fluidization velocity which is a very important parameter in the design of Fluidized bed reactors.

2.10.2.2 Pressure drop

The pressure drop through the bed is another important parameter which controls the channel and slug formation and thereby mixing of the bed material with the fluidizing fluid. At low flow rates in the packed bed, the pressure drop is approximately proportional to gas velocity up to the minimum fluidization condition(U_{mf}). With a further increase in gas velocity, the packed bed suddenly unlocks (at the onset of minimum fluidization condition), resulting in a decrease in pressure drop. With gas velocities beyond minimum fluidization(U_{mb}), the bed expands and gas bubbles are seen to rise resulting in non-homogeneity in the bed. With the increase in gas flow, the pressure drop should remain unchanged but due to bubbling and slugging (Figure 2.20) there is always a fluctuation in the pressure drop and it increases slightly. Particularly for coarse particles, the mean total pressure drop across a slugging bed may continue to increase with gas velocity higher than at the minimum fluidization condition. For fine particles the variation of

pressure drop with gas velocity is quasi-linear when they are fluidized by classical method of fluidization as they form cranks and channel(Uch). Particulate fluidization generally gives rise to a homogeneous fluidization. However, this ideal situation is not realized in practice and significant deviations have been observed (Sahoo, 2012).



Source: (Bhattacharyya, 2013)

Figure 2.20: Pressure drop across a fluidized bed

2.10.2.3 Bed expansion ratio (R)

This term is used to describe the characteristics of bed height during fluidization. This is quantitatively defined as the ratio of average height of a fluidized bed(H_{avg}) to the initial static bed height(H_{static}) at a particular flow rate of the fluidizing medium above the minimum fluidizing velocity.

Average bed height is the arithmetic mean of highest(H_{max}) and lowest(H_{min}) level occupied by top of the fluidized bed. It is denoted by "R".

$$\boldsymbol{R} = \left(\frac{\boldsymbol{H_{avg}}}{\boldsymbol{H_{static}}}\right) = \left(\frac{\boldsymbol{H_{max}+H_{min}}}{\boldsymbol{2Hs}}\right) \quad 2.6$$

The term used in the spouted bed having the same meaning at fluidization condition. It is an important parameter for fixing the height of fluidized bed required for a particular service (Sahoo, 2012). The expansion ratio of a fluidized bed depends on excess gas velocity, particle size (dp), and initial bed height (Hs).

2.10.2.4 Bed fluctuation ratio (r)

The term fluctuation ratio is used to describe the characteristics of the bed height during fluidization. This is quantitatively defined as the ratio of the highest(H_{max}) and lowest(H_{min}) levels which the top of the bed occupies for any particular gas flow rate. It is denoted by "r".

$$\boldsymbol{r} = \left(\frac{\boldsymbol{H_{max}}}{\boldsymbol{H_{min}}}\right) \quad 2.7$$

Bed fluctuation ratio has been widely used to quantify fluidization quality. So that it is also called fluidization quality. A lower value of fluctuation ratio is indicative of improved fluidization quality with less fluctuation at the top of the bed in fluidized condition (Sahoo, 2012).

2.10.2.5 Fluidization quality (FQ):

Fluidization quality is defined as the dimensionless pressure drop through the bed (Sahoo, 2012 and Suleiman et al., 2013).

$$\boldsymbol{FQ} = \frac{\boldsymbol{\Delta P.A}}{\boldsymbol{W}} \quad 2.8$$

Where FQ= Fluidisation Quality, ΔP = Pressure Differential (atm), A = Area (m^2) and

W = Weight of particulate (kg)

A FQ equal to unity characterizes a homogeneous fluidization of either individual particles or ephemeral agglomerates. In contrast, channeling and slugging phenomena lead to poorer fluidization qualities (Sahoo, 2012).

The solid substrate (the catalytic material upon which chemical species react) material in the fluidized bed reactor is typically supported by a porous plate, known as a distributor. The fluid is then forced through the distributor up through the solid material. At lower fluid velocities, the solids remain in place as the fluid passes through the voids in the material. This is known as a packed reactor. As the fluid velocity is increased, the reactor will reach a stage where the force of the fluid on the solids is enough to balance the weight of the solid material. This stage is known as incipient fluidization and occurs at minimum fluidization velocity. Once this minimum velocity is surpassed the content of the reactor bed begin to expand and swirl around much like an agitated tank or boiling pot of water, the reactor is now fluidized bed. Depending on the operating conditions and properties of solid phase various flow regime can be observed in the reactor, example of flow regime are fixed bed, particulate, bubbling, slug flow and turbulent regimes.

2.10.3 Mass of solid in the bed

There is a drag force exerted on the solid particles by the flowing fluid, and at low fluid velocities the pressure drop resulting from this drag follows the Ergun equation, (Brown and Fogler, 2008), just as for any other type of packed bed. When the fluid velocity is increased to a

certain value however, the total drag on the particles will equal the weight of the bed, and the particles will begin to lift and barely fluidize. This is expressed in equation 2.9;

$$\boldsymbol{W_s = \rho_s A_s h(1-\varepsilon)} \qquad 2.9$$

Where, Ws =Mass of solids in the bed,

ρ_s = Density of solid (kg/m^3),

A_s = Cross-sectional area of solid (m^2),

h = Height of the bed settled before the particles start to lift (m), and

ϵ = Void fraction of bed.

The void fraction of bed is expressed as;

$$\boldsymbol{\varepsilon = 1 - \frac{Mp}{\rho_s \times Vt}} \qquad 2.10$$

Where Mp is mass of particle and Vt is total volume.

The macroscopic observables in fluidized beds are the fluid pressure drop (ΔP) needed to cause the fluid to flow through the bed of solids, the fluid velocity (u), and the density of solids (ρ_s). The Ergun equation (Idris et al., 2007) is represented as in equation 2.11;

$$\boldsymbol{\frac{\Delta p}{h} = 150 \frac{(1-\varepsilon)^2}{\varepsilon^3} \frac{\mu u}{d_p^2} + 1.75 \frac{(1-\varepsilon)}{\varepsilon^3} \frac{u \rho_f}{d_p}} \qquad 2.11$$

Where, ΔP = pressure drop (atm)

h = Height of the bed (m),

μ = fluid viscosity (kg/ms),

ρ_f = Fluid density (kg/m^3),

ϵ = Void fraction of bed,

u = fluid velocity (m/s), and

d_p = Particle diameter (m).

The equation for turbulent flow is different from streamline situation. The particles in the bed will remain in a parked bed as the gravitational forces holding the solid particles down are greater than the force exerted by the fluid flowing up through the bed particles. At the point where the two forces become equal, the solid particles begin to move up. The force balance describes this condition known as incipient fluidization (Idris et al., 2007 & Brown and Fogler, 2008) is given in equation 2.12;

$$\frac{\Delta p}{h} = (\rho_s - \rho_f)(1 - \varepsilon)g \qquad 2.12$$

Where ρ_s = Density of solid (kg/m^3), ε = Void fraction, g = acceleration due to gravity (m/s^2)

The fluidization at minimum velocity, u_{mf} in terms of parameters for the fluid, solid and bed is expressed (Missen et al., 1999) as in equation 2.13;

$$u_{mf}^2 + \frac{150(1-\varepsilon_{em})\mu_f}{1.75\rho_f d_p^l} u_{mf} - \frac{(\rho_p - \rho_f)\varepsilon_{mf}^3 d_p^l}{1.75\rho_f} = 0 \qquad 2.13$$

For relative small particles and small Reynolds number the minimum fluidized velocity (u_{mf}) is expressed as in equation 2.14;

$$u_{mf} = \frac{8(\rho_p - \rho_f)(d_p^l)^2}{150(1-\varepsilon_{mf}) \div \varepsilon_{mf}^3} \qquad 2.14$$

Where $\boldsymbol{\varepsilon_{mf}}$ = Void fraction at minimum fluidization.

The minimum fluidization velocity (u_{mf}) for relative large particles and large Reynolds number is given by equation 2.15;

$$\boldsymbol{u_{mf}} = \sqrt[2]{\left[\frac{g(\rho_p - \rho_f)\varepsilon_{mf}^3 d_p^{|}}{1{,}75\rho_f}\right]} \qquad 2.15$$

Where:

Umf = Minimum fluidization velocity (m/s)

g = Acceleration due to gravity (m/s^2),

ρ_f = Fluid density (kg/m^3),

ρ_d = Particle density (kg/m^3),

ϵ_{mf} = Void fraction of bed at minimum fluidization,

d_p = Particle diameter (m)

2.10.4 Terminal velocity, Ut

The solid particles are entrained when the upward velocity of fluid is sufficiently high and thus they are carried up with the fluid. At this point elutriation occurs. This velocity is called elutriation velocity. Elutriation is the selective removal of solid particles by entrainment on the basis of size (Missen et al., 1999). The minimum elutriation velocity for particles of a given size is the velocity at incipient entrainment, and is assumed to be equal to the terminal velocity, ut. The terminal velocity for spherical particles at small Re is expressed (Missen et al., 1999) in equation 2.16 as;

$$\boldsymbol{u_t = \frac{g(\rho_p - \rho_f)d_p^2}{18\mu_f}} \quad 2.16$$

Proper fluidization occurs at a velocity called actual fluidized velocity u_f (Hartog et al, 2008 and Missen et al, 1999) as shown in figure 2.20 the relationship between the minimum fluidized velocity u_{mf} and terminal velocity u_t is given in equation 2.17;

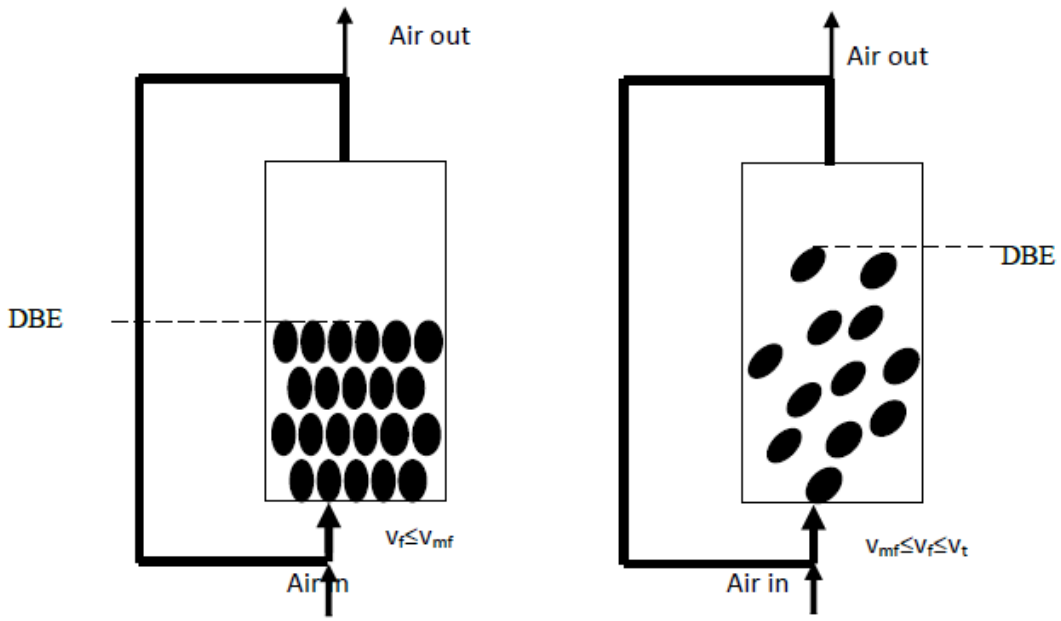
$$\boldsymbol{u_{mf} < u_f < u_t} \quad 2.17$$

If the gas velocity is increased to a sufficiently high value, however, the drag on an individual particle will surpass the gravitational force on the particle, and the particle will be entrained in a gas and carried out of the bed. The point at which the drag on an individual particle is about to exceed the gravitational force exerted on it is called the maximum fluidization velocity(Suleiman et al 2013).

The fluidized velocity (u_f) expressed by Kozeny-Carmen equation is as in equation 2.18

$$\boldsymbol{u_f = \frac{(\rho_p - \rho_f)g d_p^2}{150\mu} \frac{\varepsilon^3}{1-\varepsilon}} \quad 2.18$$

Figure 2.20 shows diagrammatic representation of conceptual fluidized reactor containing uniform particles with degree of bed expansion (DBE) as related to flow velocity V_f (Suleiman et al, 2013).



Source: (Suleiman et al, 2013)

Figure 2.21: Conceptual Fluidized Reactor

2.10.5 Reactor design

An industrial chemical reactor is a complex device in which heat transfer, mass transfer, diffusion and friction must be considered and it must be safe and controllable (Sanju, 2010). In large vessels, problem of mixing of reactants, flow distribution, residence time distribution and efficient utilization of the surface of porous catalyst also arise. An economic balance of these factors result to a successful commercial unit (Sanju 2010).

A general procedure for reactor design is as follows:

1. The kinetic and thermodynamic data on the desired reaction is initially collected either from literature or from an existing process. Values will be needed for the rate of reaction over a range of operating conditions, for example, pressure, temperature, catalyst concentration and flowrate. This data may be normally obtained from either laboratory or pilot plant studies.

2. Data on physical properties is required for the design of the reactor. This may be either estimated, or collected from the literature or obtained by taking laboratory measurements.

3. The rate controlling mechanism which has a predominant role is then identified, for example, kinetic, mass or heat transfer.

4. A suitable reactor type is then chosen, based on experience with similar studies or from the laboratory and pilot plant work.

5. Selection of optimal reaction conditions is initially made in order to obtain the desired yield

6. The size of the reactor is decided and its performance estimated. Since exact analytical solutions of the design relationship are rarely possible, semi empirical methods based on the analysis of idealized reactors are used.

7. Materials for the construction of the reactor is/are selected.

8. A preliminary mechanical design for the reactor including the vessel design, heat transfer surfaces etc., is made.

9. The design is optimized and validated

10. An approximate cost of the proposed and validated design is then calculated.

In choosing the reactor conditions, and optimizing the design, the interaction of the reactor design with the other process operations must not be overlooked. The degree of conversion of raw materials in the reactor will determine the size and the cost of any equipment needed to separate and recycle unreacted materials. In these circumstances the reactor and associated equipment must be optimized as a unit (Sanju, 2010).

2.11 Process Selection

The design of a process plant involves a sequence of process selection, process description, development of process flow diagram, material and energy balances, detailed equipment design and equipment specification. All these will be shown in this section.

2.11.1 Types of FCC designs

There are a number of different proprietary designs that have been developed for modern FCC units. Each design is available under a license that must be purchased from the design developer by any petroleum refining company desiring to construct and operate an FCC of a given design. Basically, there are two different configurations for an FCC unit: the "stacked" type where the reactor and the catalyst regenerator are contained in a single vessel with the reactor above the catalyst regenerator and the "side-by-side" type where the reactor and catalyst regenerator are in two separate vessels (Ibsen, 2006). Several designs are available but these are the major FCC designers and licensors:

2.11.1.1 Side-by-side configuration:

The following Licensors used the side by side configuration;

i- Chicago Bridge and Iron Company (CB&I)

ii- ExxonMobil Research and Engineering (EMRE)

iii- Shell Global Solutions International

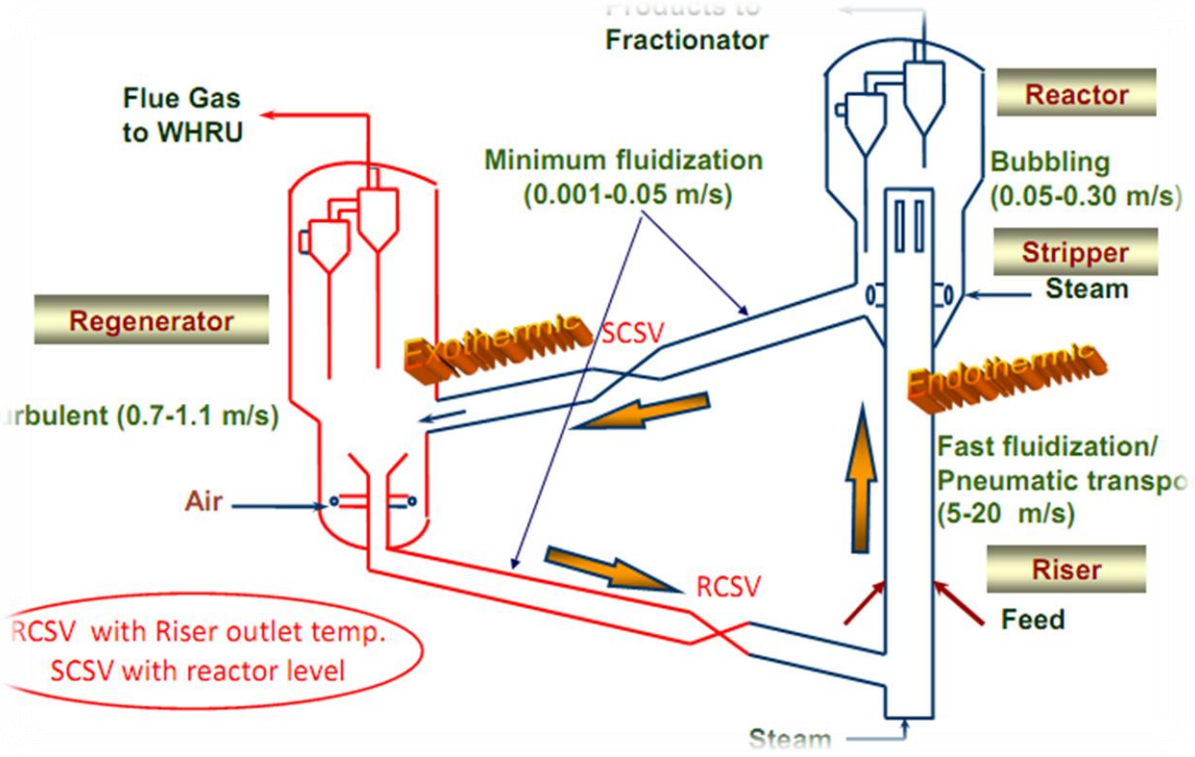
iv- Stone & Webster Engineering Corporation (The Shaw Group) / Institute François Petrol (IFP)

v- Universal Oil Products (UOP) — currently fully owned subsidiary of Honeywell

2.11.1.2 Stacked configuration:

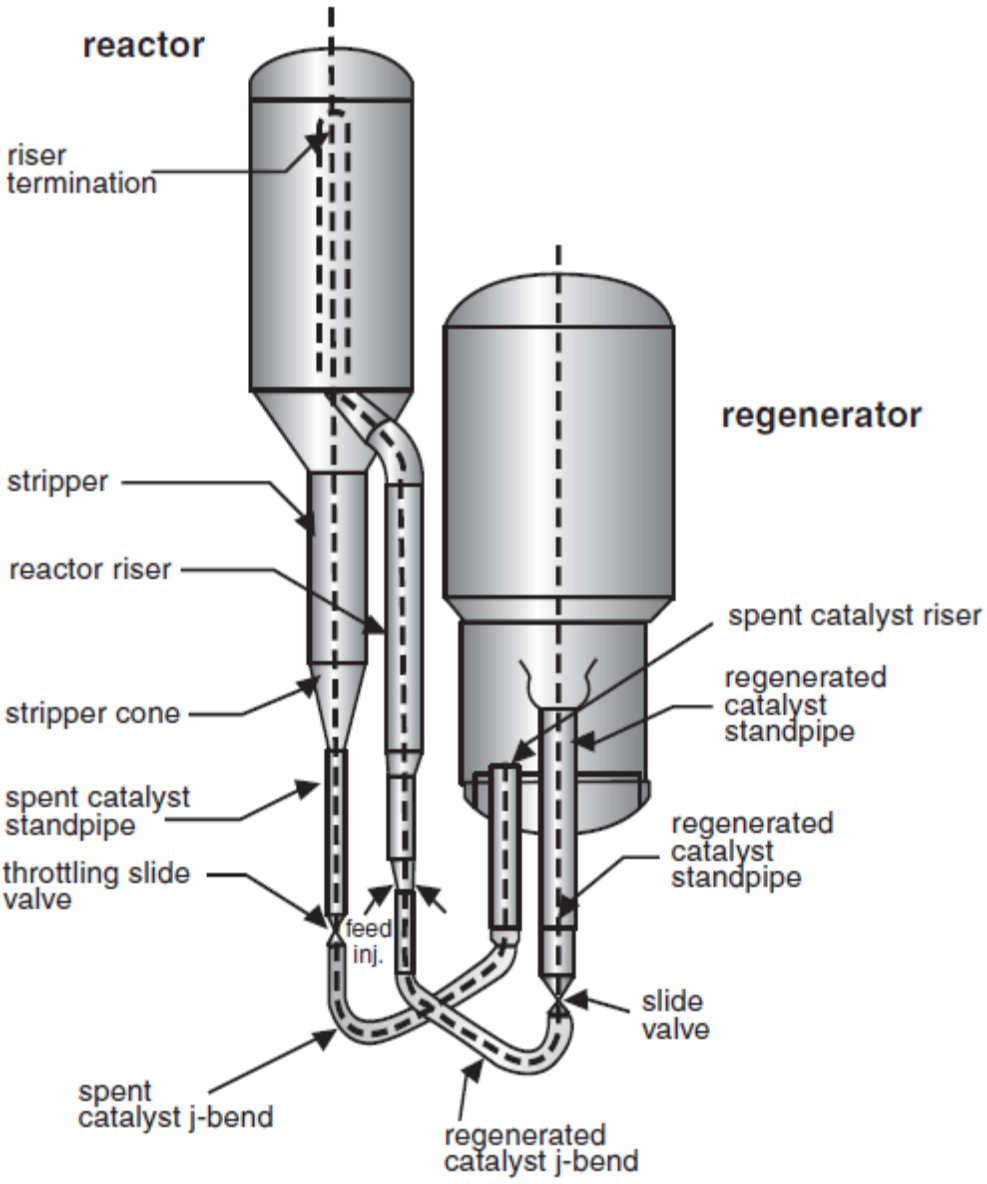
Kellogg Brown & Root (KBR)

All the proprietary design licensors claim to have unique features and advantages. It will be pertinent to mention that all the licensors have designed and constructed FCC units that have operated quite satisfactorily. All other feed pre-treatment and product separation processes remain the same with difference only in the reactor design as shown diagrammatically in Figure 2.22 – 2.27.



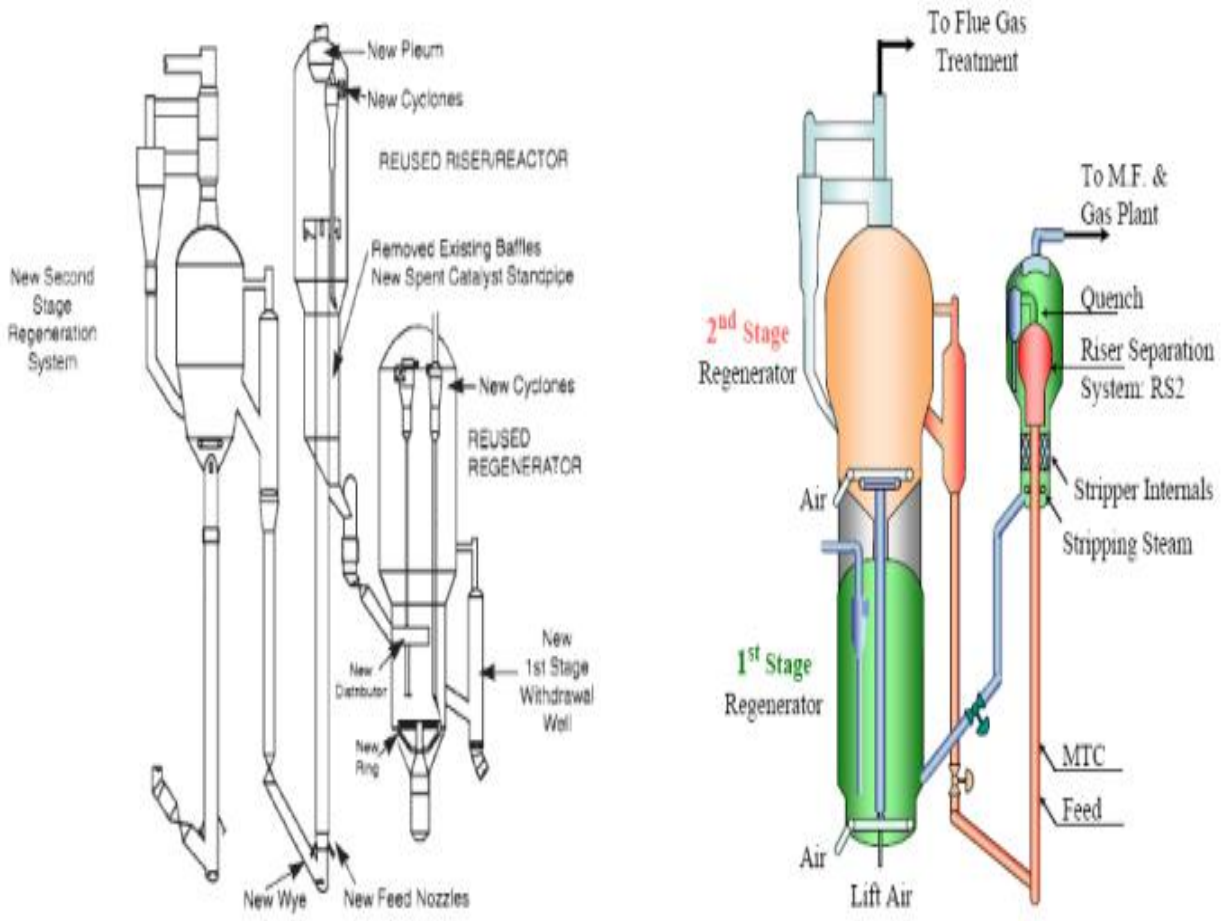
Source: (Bhattacharyya, 2013)

Figure 2.22: Typical Side-by-Side FCC Reactor



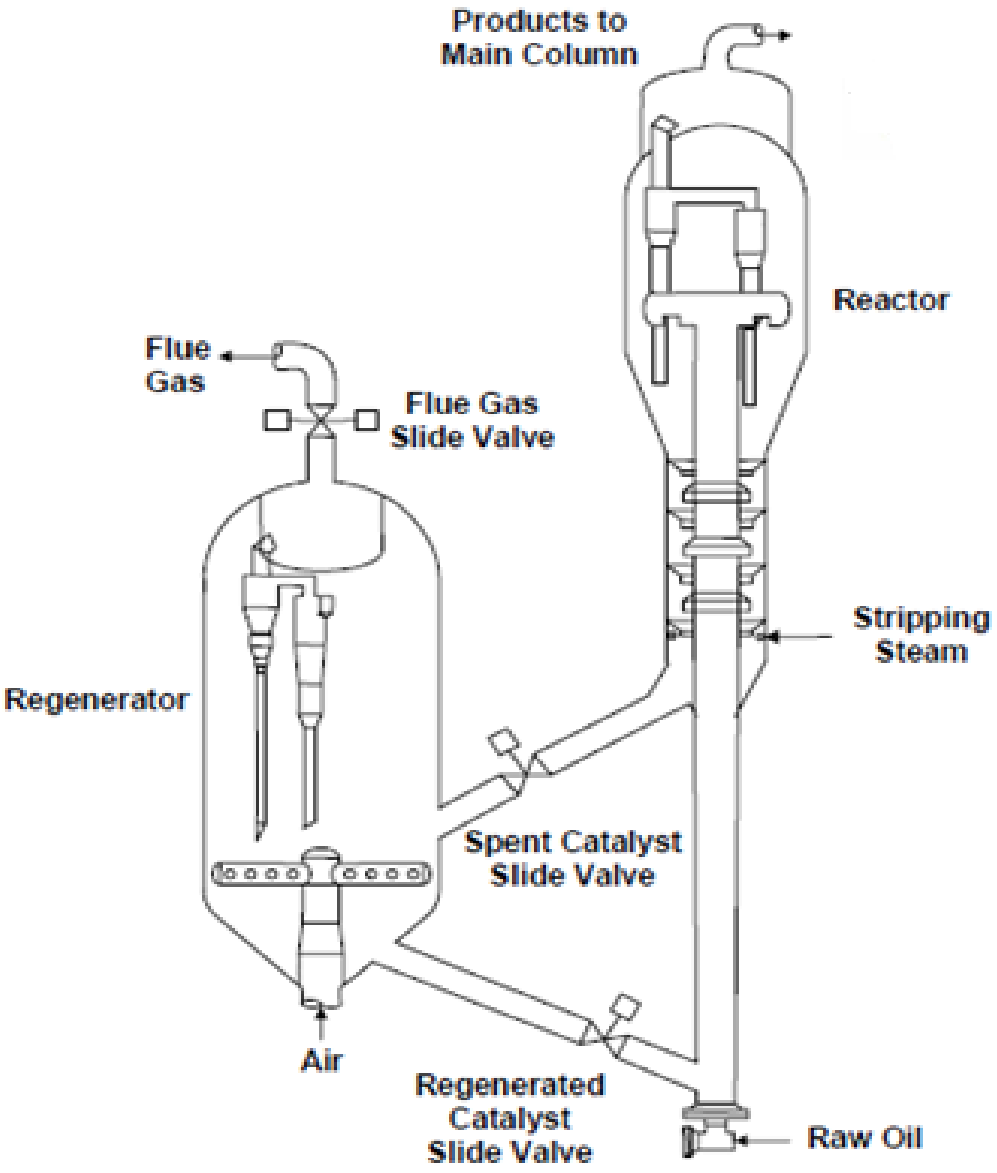
Source: (Bhattacharyya, 2013)

Figure 2.23: ExxonMobil Flexi-cracker side-by-side design



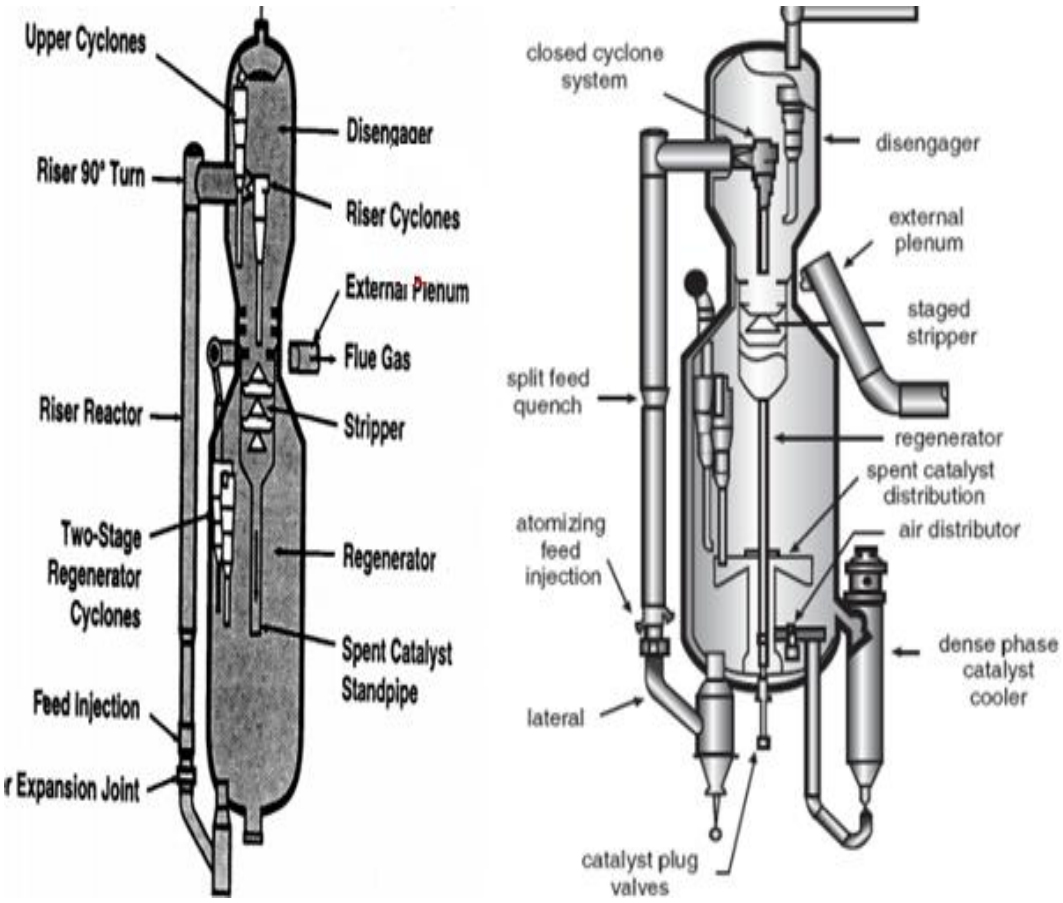
Source: (Bhattacharyya, 2013)

Figure 2.24: FCC side-by-side Reactor Design (a) RFCC Unit by SWEC (b) R2R Unit by Axens



(Bhattacharyya, 2013)

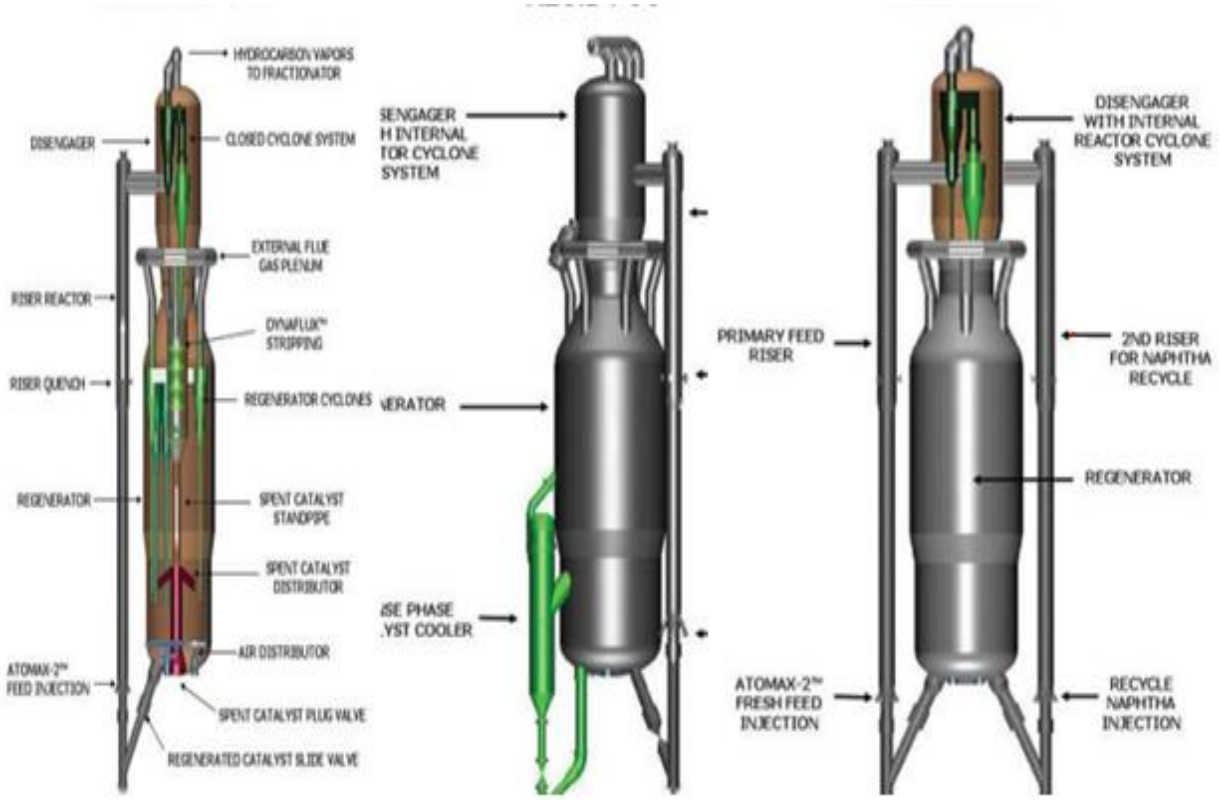
Figure 2.25: UOP Side-by-Side Design



(a) (b)

Source: (Bhattacharyya, 2013)

Figure 2.26: Kellogg Orthoflow Stacked Design (a) FCC Converter (b) Resid FCC Converter



(a) (b) (c)

Source: (Bhattacharyya, 2013)

Figure 2.27: Other Stacked designs by Kellogg (a) Orthoflow^{TM} (b) Resid FCC (c) Maxifin^{TM}

CHAPTER THREE

METHODOLOGY

3.1 Preamble

This chapter presents a step by step description of activities carried out in the design and fabrication of 0.56bbl/hr Fluid Catalytic Cracking Unit. These stages are itemized in a block diagram shown in figure 3.1;

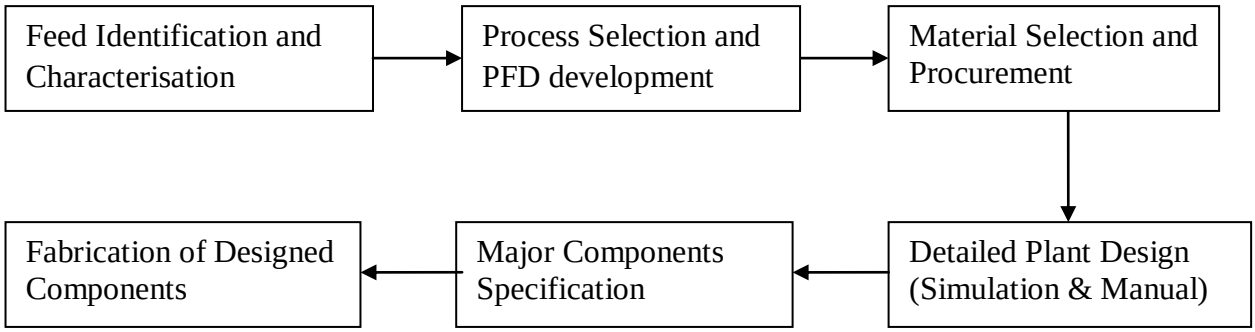


Figure 3.1: Methodology block diagram

3.2 Design Basis

The identity of the feed, its properties and characteristics were first of all established before a design basis was chosen. The design basis, feedstock composition and properties of Mini FCCU is presented in Tables 3.1, 3.2 and 3.3

Table 3.1:Mini FCCU Design Basis

Mini FCCU Capacity	5bbl/batch (batch time is 9hrs)
Feed Type	Escravos VGO/HGO/Waxy Distillate
Source	Atmospheric/Vacuum Distillation Unit
Specific Gravity	0.890 Kg/m^3

Table 3.2 Feed Composition

Components	Mol. Wt. (Kg/Kmol)	Weight % (Kg/batch)	Weight % (Kg/hr)
VGO	214	0.4264	0.0474
HGO	265	0.3831	0.0426
W/Distillate	266	0.1904	0.0212

Table 3.3: Typical KRPC FCC feed charactristics

Properties	**FCC Specification**
S.G 15/4^{0}C	0.890
Nitrogen ppm	590
Viscosity Cst	190
Sulphur wt%	0.1
CCR wt%	1
Asphaltene wt%	0
V wt ppm	0
Ni wt ppm	0.2
DISTILLATION	
IBP ^{0}C	284
10%	351
30%	380
50%	409
70%	445
90%	490
K factor	12
Mol wt	371

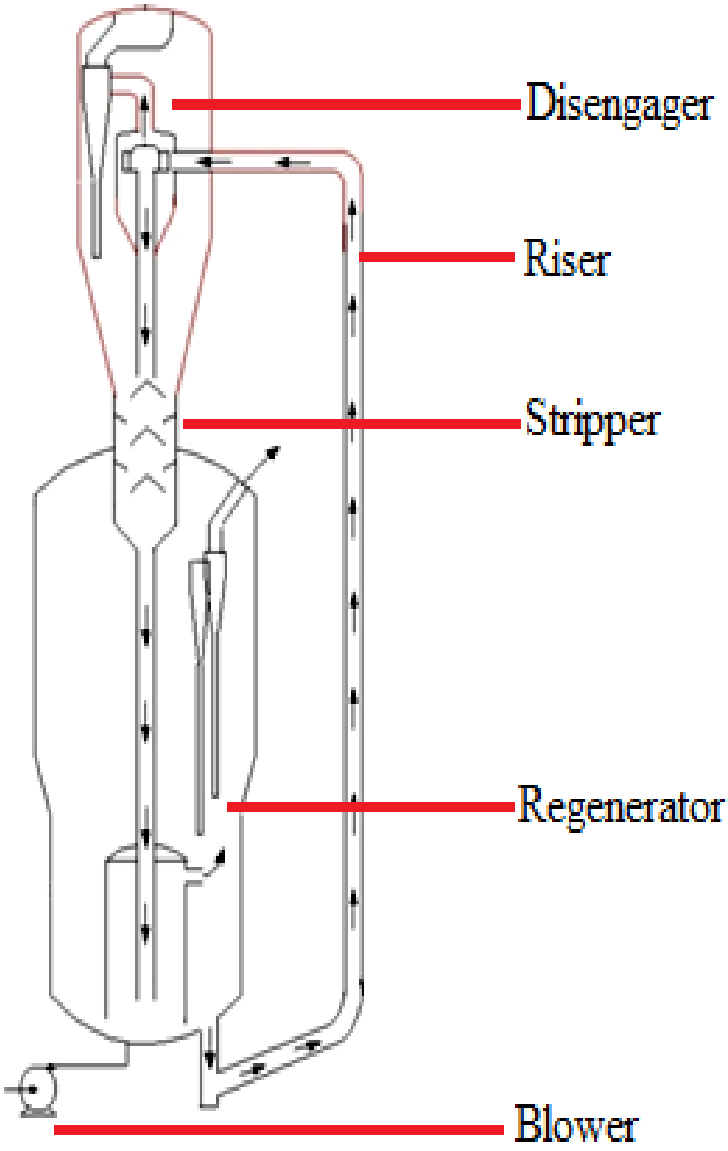
Source:(KRPC, 1980)

3.3 Process Selection and PFD Development

The stack process configuration was adopted in this design due to the design capacity of the proposed FCC unit. This design method is suitable for small capacity design (Ibsen, 2006) and

hence it will be applied here. The stack FCC reactor and the whole process is depicted in a simplified process flow diagram as shown in Figure 3.7 and 3.8 and a list of equipments and their respective codes as depicted in the process flow diagram is as follows:

1. Feed Surge Drum (D-01)
2. Feed/LCO Heat Exchanger (E-01)
3. Feed/DCO Heat Exchanger (E-02)
4. Feed/Slurry Heat Exchanger (E-03)
5. Feed Fired Heater (F-01)
6. FCC Reactor (R-01)
7. Main Fractionator (C-01)
8. Overhead Condenser (A-01)
9. Overhead Separator (D-02)
10. LCO Side Stripper (C-02)
11. DCO Side Stripper (C-03)



Source: (Ibsen, 2006)

Figure 3.1 Simplified Stack FCC reactor

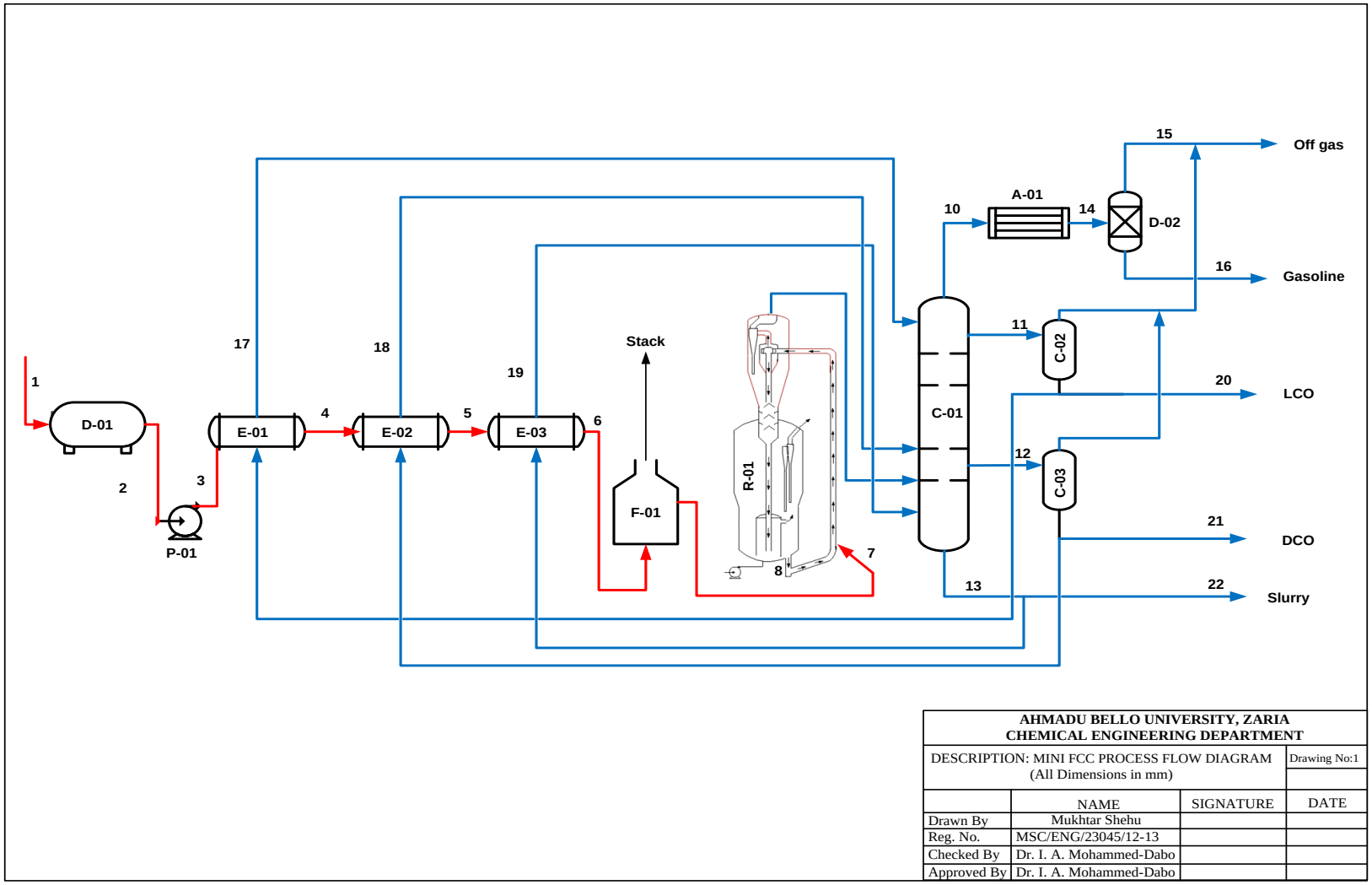


Figure 3.2: Process Flow Diagram of Fluid Catalytic Cracking

3.4 Process Description

Heavy gas oils, light vacuum gas oil and heavy vacuum gas oil are the feed streams to fluid catalytic cracking unit. The FCCU breaks up the heavy oils into more volatile gasoline materials. The feed is introduced into the unit via the surge drum D01 then it passes through a series of heat exchangers E01 (light cycle oil), E02 (heavy cycle oil) and E03 (Decanted oil) where it is heated sequencially from (26-90)^{o}C,(90-158)^{o}C and (158-220)^{o}C respectively via a feed pump P01, and passes to the fresh feed furnace F01 to raise the temperature further to the required temperature of 320^{0}C. It is sent to the converter R01 for conversion into light products. FCCU oil feed is atomized into a stream of very hot regenerated catalyst at the bottom of a vertical "riser" reactor R01. The charge is combined with a recycle stream within the riser, vaporized, and raised to reactor temperature (520 - 560° C) by the hot catalyst. As the mixture travels up the riser, the charge is cracked at 10-30 psi, the mixture of catalyst and now-vaporized oil shoots to the top of the riser (disengager)(Figure 3.7) where catalyst and hydrocarbon reaction products are separated by inertial means followed by passage of the vapour product through a cyclone separator(Figure 3.7) to remove virtually all entrained catalyst powder.

During its passage through the riser most of the FCCU feed is broken down into light olefins, FCC gasoline, a diesel range stream (light cycle oil), and small amount of heavier oil (decanted oil) which can be blended into bunker or other heavy fuel oil. However, a small amount of the feedstock breaks down to coke which is deposited on the catalyst particle surfaces. Spent catalyst from the reaction disengaging vessel (Figure 3.7) flows down through a steam-stripping chamber where any remaining volatile hydrocarbon is stripped out and joins the riser product stream. Spent catalyst then flows to the Regenerator, where the coke is burned off using air supplied by an air blower and uniformly distributed at the bottom of the Regenerator by an air grid. The upward velocity of the air tends to lift the catalyst particles, so that the bed of catalyst behaves as

a fluid. Flue gas from the regenerator is sent to the CO boiler via a slight valve and an orifice chamber.

Reaction product vapour from the overhead of the disengager is processed at low pressure in the Main Fractionating Column C01 where decanted oil and light cycle oil streams are separated stripped in side strippers C02 and C03 partly recycled as pump around back to the column C01 via heat exchange in feed preheaters E01 and E02 and partly sent to tank. The bottom slurry is partly recycled via heat exchange in E03 back to the column C01 and partly sent to tank. The overhead vapour from the main Fractionator is further condensed to liquid in the overhead condenser A01 and further separated in high pressure separator D02. Part of the condensed gasoline is recycled as reflux to the column and partly sent to storage while the off gas is sent to the vapour recovery unit (VRU). Figure 3.8 depicts a simplified PFD and plate IV a-c depict a 3D diagram of the entire process plant.

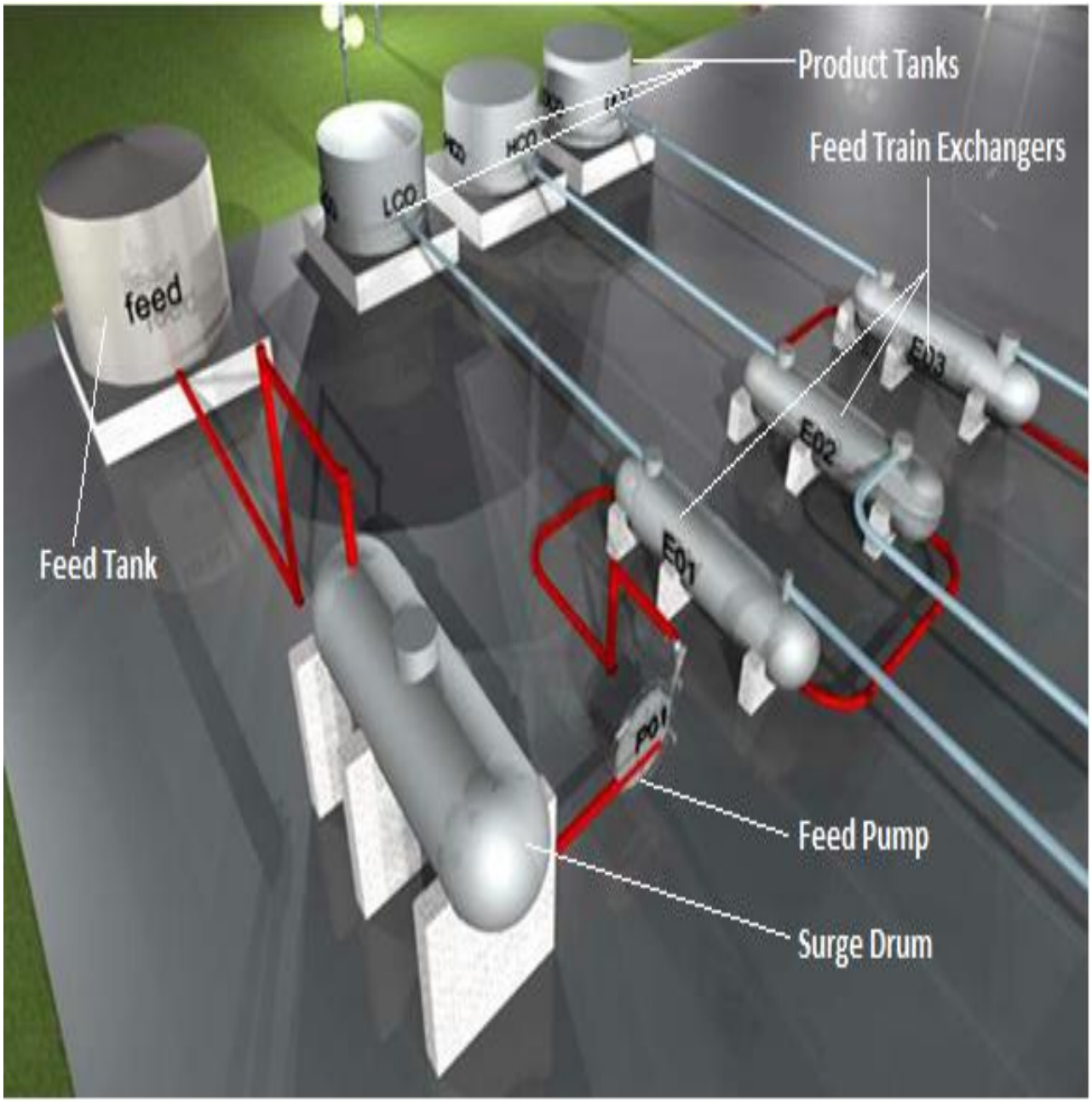


Plate IV a: Dissected View of FCCU Pilot Plant 3D Diagram

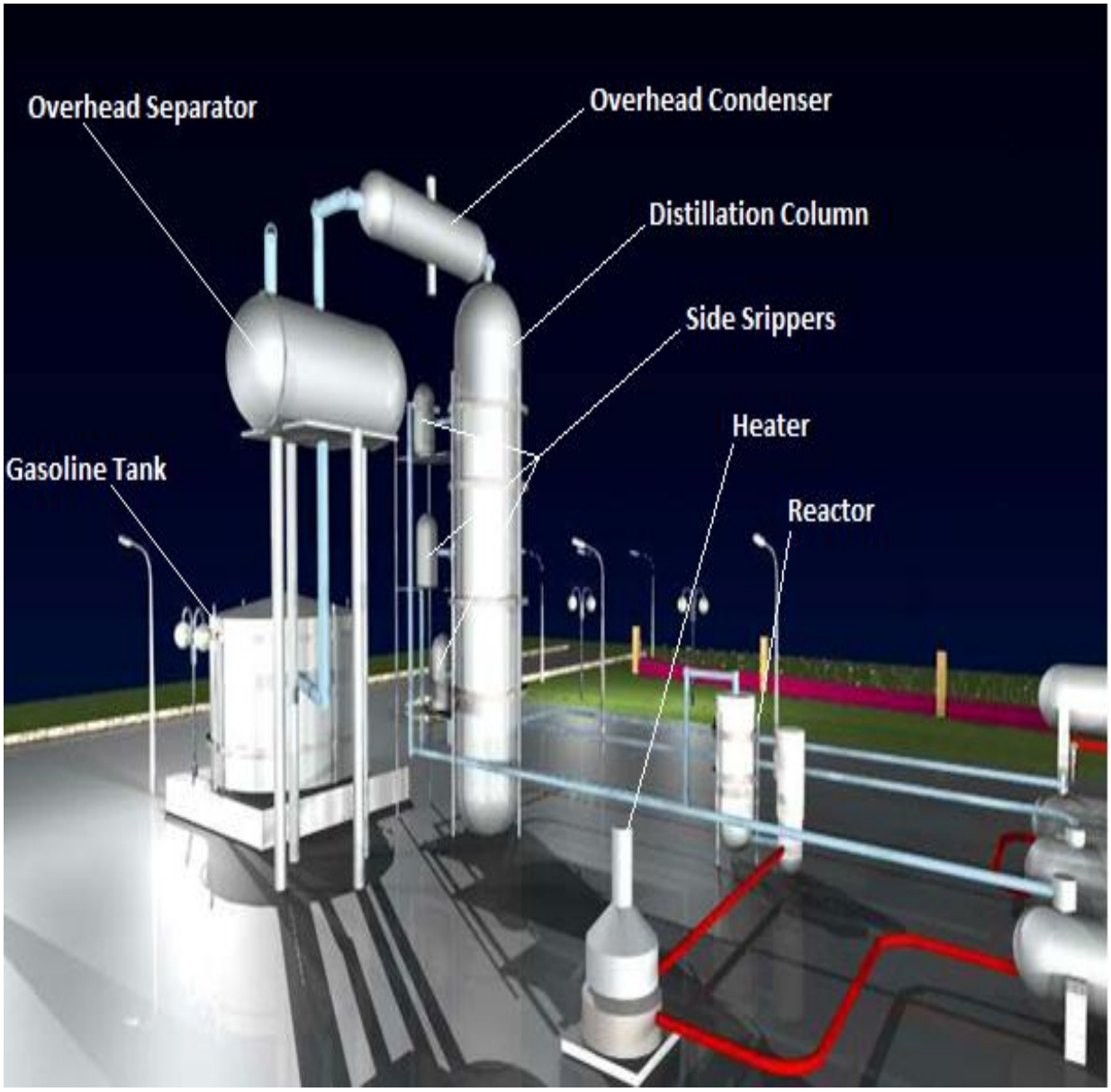


Plate IV b: Dissected View of FCCU Pilot Plant 3D Diagram

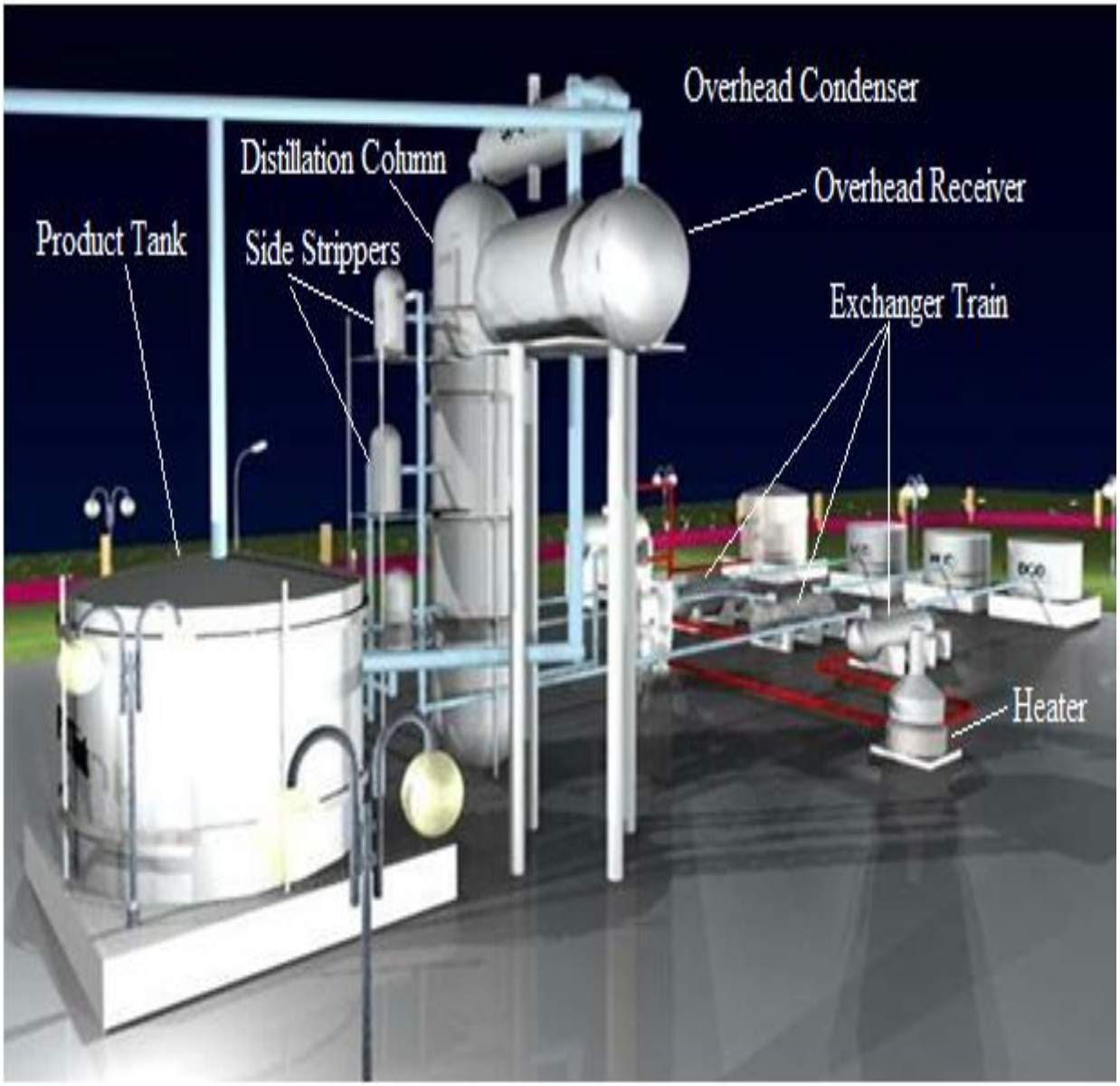


Plate IV c: Complete View of FCCU Pilot Plant 3D Diagram

3.5 Manual Design Procedure

A basis of 5barrels/batch (0.56barrels/hr) was selected for the design and all assumptions were clearely stated as can be seen in the appendix. Figure 3.3 presents the Mini FCCU Process Flow Diagram showing streams used for the material and energy balances. All assumptions on each equipment were clearly stated in the appendix C while inlet and outlet streams were clearly depicted in Figures 3.4 – 3.13

3.5 .1 Material and energy balances

Material and energy balances were carried out manually and the summary of the balances is shown in the result section while the detail is shown in appendices A&B. A basis of 5bbl/ batch (0.56barrels/hr) was selected for the calculations. Mathcad was used as a manual numerical calculator for simplicity and accuracy.

3.5.2 Detailed equipment design

Design of individual equipments were carried out using corresponding empherical correlations from the literature. The equipments designed include the surge drum, heat exchangers, fired heater, FCC reactor, side strippers, overhead condenser and overhead receiver. Material selection was also carried out and all specifications were made available for fabrication.

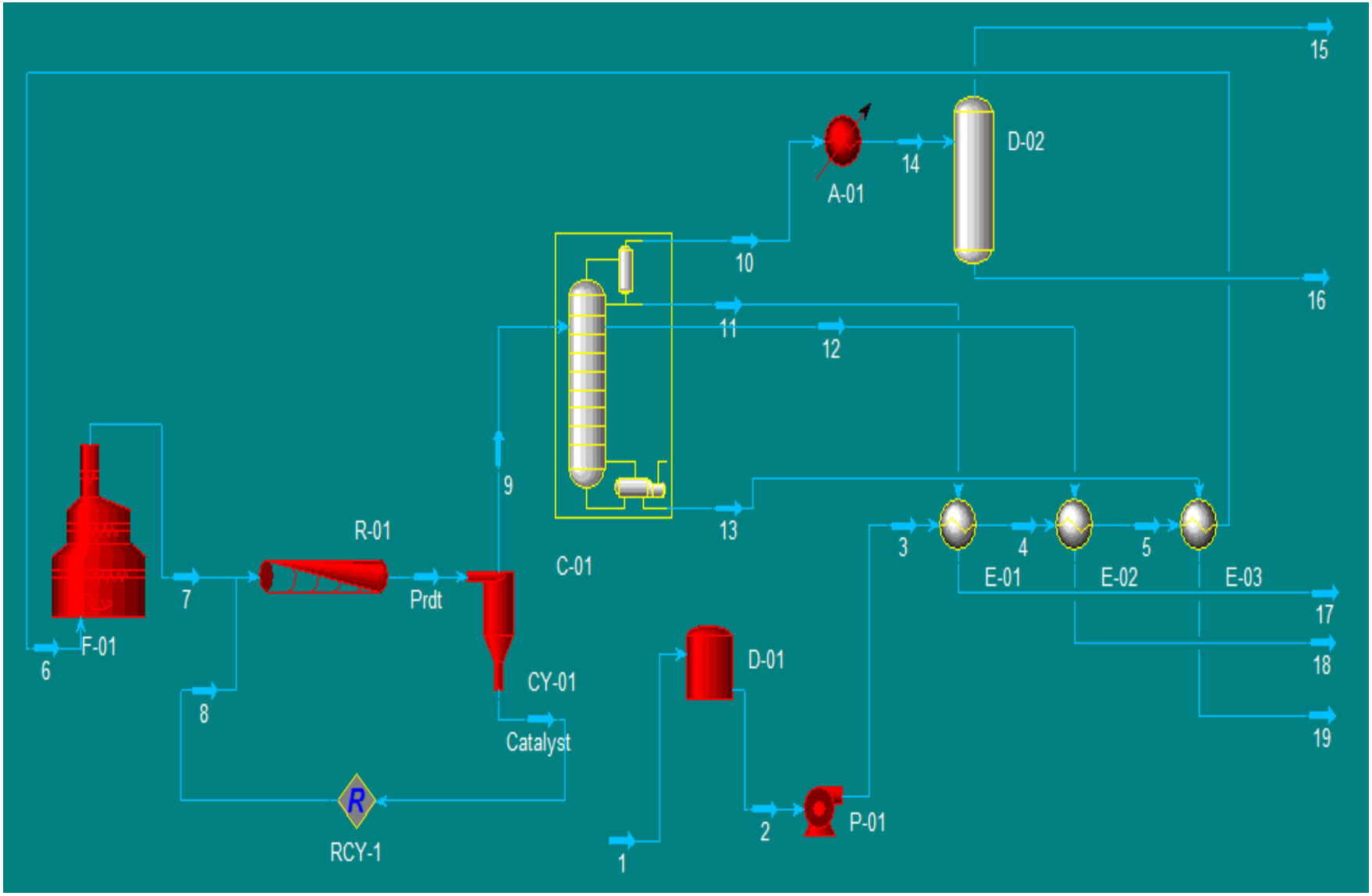


Figure 3.3: Manual Design PFD of Pilot FCC Unit

3.5.3 FCC Unit Surge Drum D-01

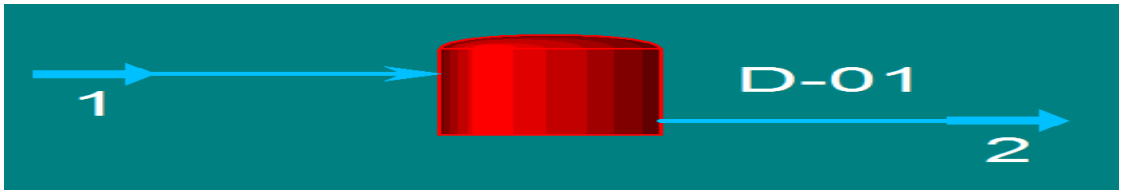


Figure 3.4 Flow Streams of Surge Drum (D-01)

3.5.4 FCC Unit Feed Surge Pump P-01



Figure 3.5: Flow Streams of Surge Pump (P-01)

3.5.5 FCC Unit Heat Exchanger E-01

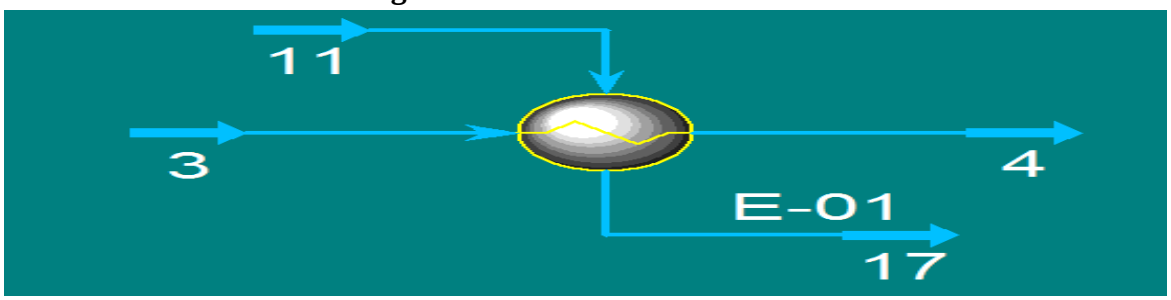


Figure 3.6: Flow Streams of FCC Unit Heat Exchanger (E-01)

3.5.6 FCC Unit Heat Exchanger E-02

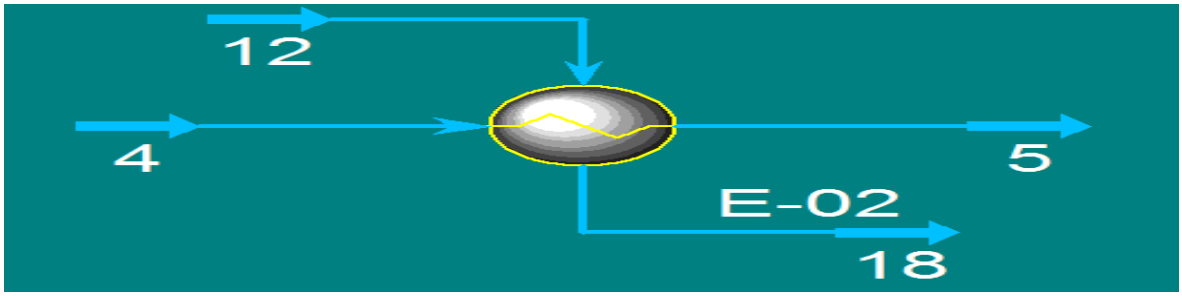


Figure 3.7: Flow Streams of FCC Unit Heat Exchanger (E-02)

3.5.7 FCC unit heat exchanger E-03

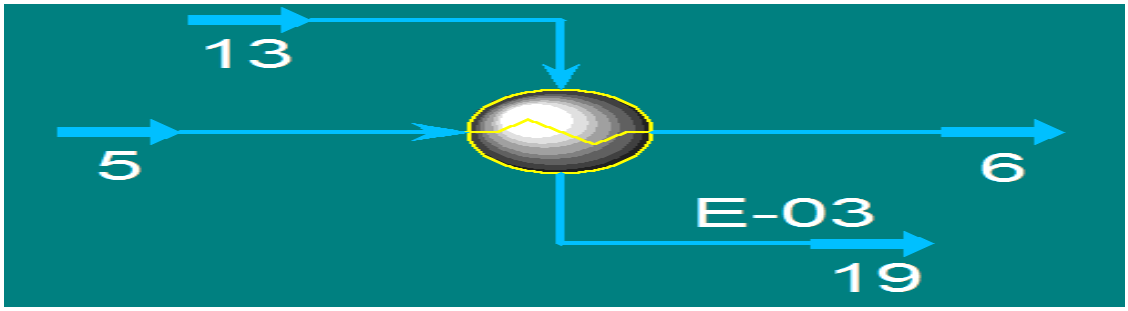


Figure 3.8: Flow Streams of FCC Unit Heat Exchanger (E-03)

3.5.8 FCC unit feed preheater F-01

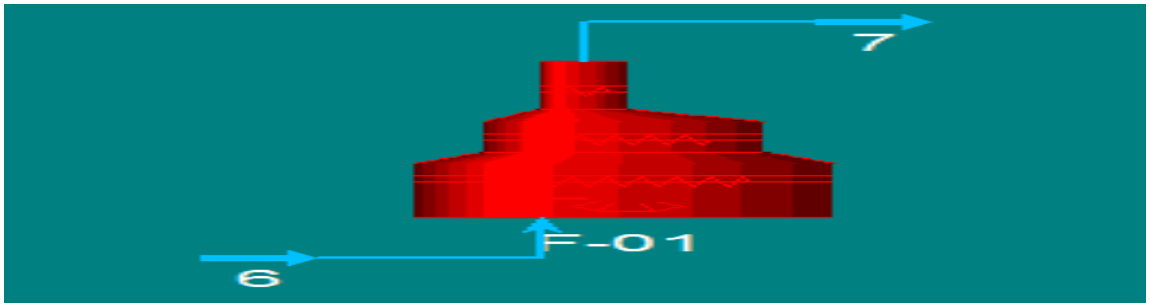


Figure 3.9: Flow Streams of FCC Unit Preheater (F-01)

3.5.9 FCC unit reactor R-01

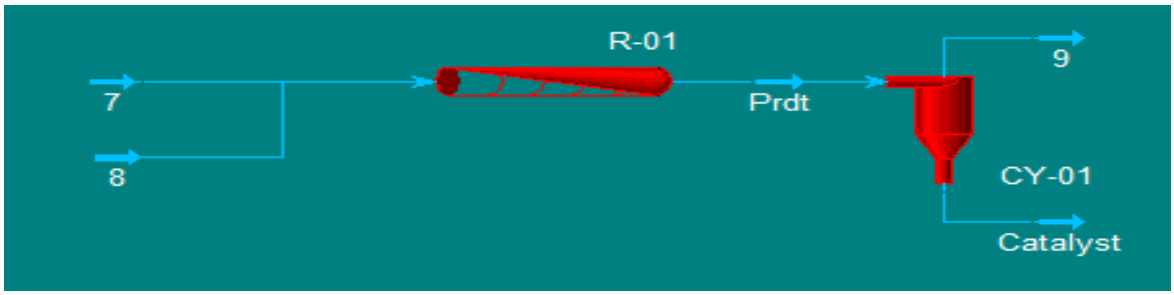


Figure 3.10: Flow Streams of FCC Unit Reactor (R-01)

3.5.10 FCC unit main fractionator C-01

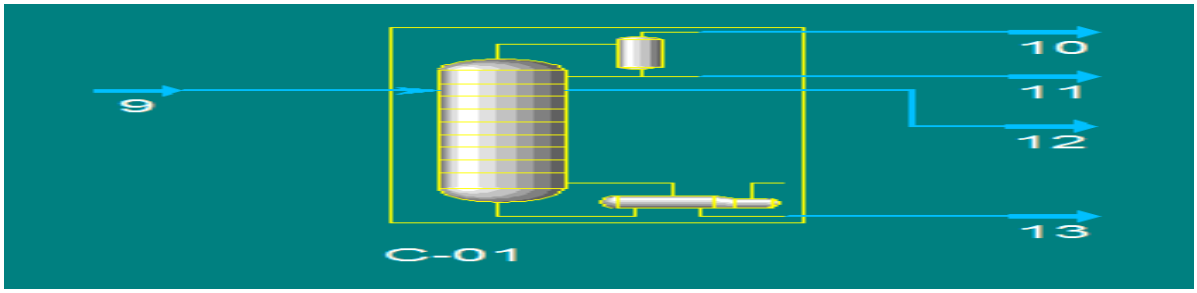


Figure 3.11: Flow Streams of FCC Unit Main Fractionator (C-01)

3.5.11 FCC unit air fin cooler A-01

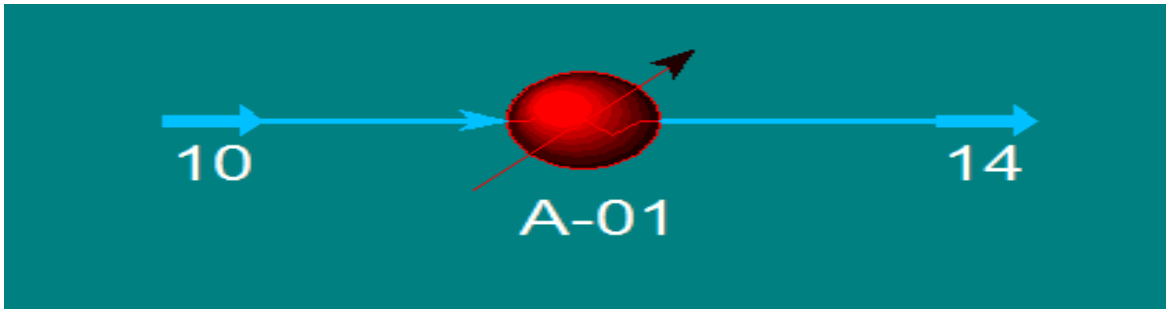


Figure 3.12: Flow Streams of FCC Unit Air Fin Cooler (A-01)

3.5.12 FCC unit overhead separator D-02

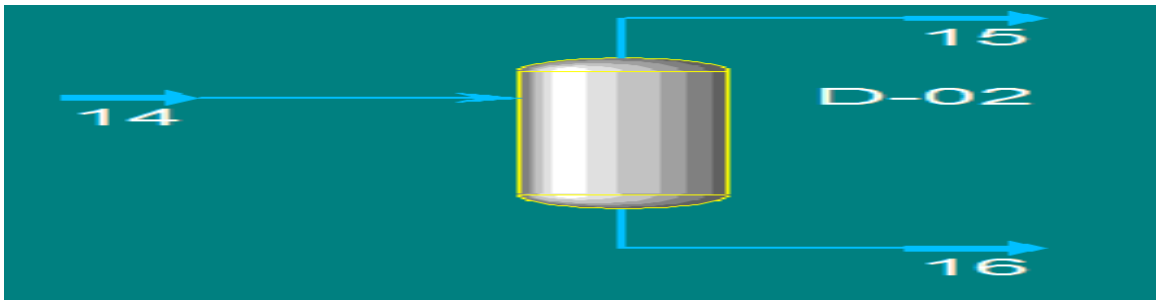


Figure 3.13: Flow Streams of FCC Unit overhead Separator (D-02)

3.6 Computer (Hysys) Simulation

Computer simulation of the proposed pilot plant was carried out using Hysys 7.2 simulation soft ware according to the following procedure.

3.6.1 Process simulation procedure

I. Data Collection: Operation data were collected from literature(Thermodynamic data in Hysys) and Kaduna Refining and Petrochemical Company (KRPC) Kaduna(Crude assay). Process flow diagram was developed for the Pilot Plant (figure 3.14) from literature (Raji et al, 2012 and Nuhu et al, 2012) and experience gained from real situation in KRPC.

II. Construction of Pilot FCC model in a Process Simulator: Building the Model of the developed Pilot FCC process flow diagram in Hysys using the data collected in 1 above as shown in Figure 3.14:

III. Computer Simulation: Computer simulation of the constructed model was carried out using Hysys 7.2 detail of which is explained in section 3.6.2.

3.6.2 Process simulation

In order to obtain reasonable design parameters for the pilot plant, computer simulation software (Hysys 7.2) was used to carry out complete modeling and process simulation of the plant. This software provides reliable information on process operation with comprehensive thermodynamic packages. Vast component libraries and advanced calculation technique are contained in the package. The process simulation software Hysys Version 7.2 was used.

Mainly, process simulation involved the following procedure:

i- Defining feed Chemical components

ii- Selecting thermodynamic package

iii- Specifying plant capacity

iv- Assigning proper operating units and Setting up input operating conditions (pressure, temperature, flowrate, catalyst information, etc).

Inbuilt data on most components such as water, hydrocarbons, CO, CO_2, NO_2, SO_2 etc are available in the Hysys component library (Raji et al, 2012). As can be seen in Figure 4.5.1 a process flow diagram of the proposed pilot plant was constructed and simulated in a Hysys environment and the result of the simulation is shown in Tables 4.2.1 - 4.2.9(result section) while the details are presented in appendix D. This result was compared with the manual design carried out to confirm the accuracy of manual calculation.

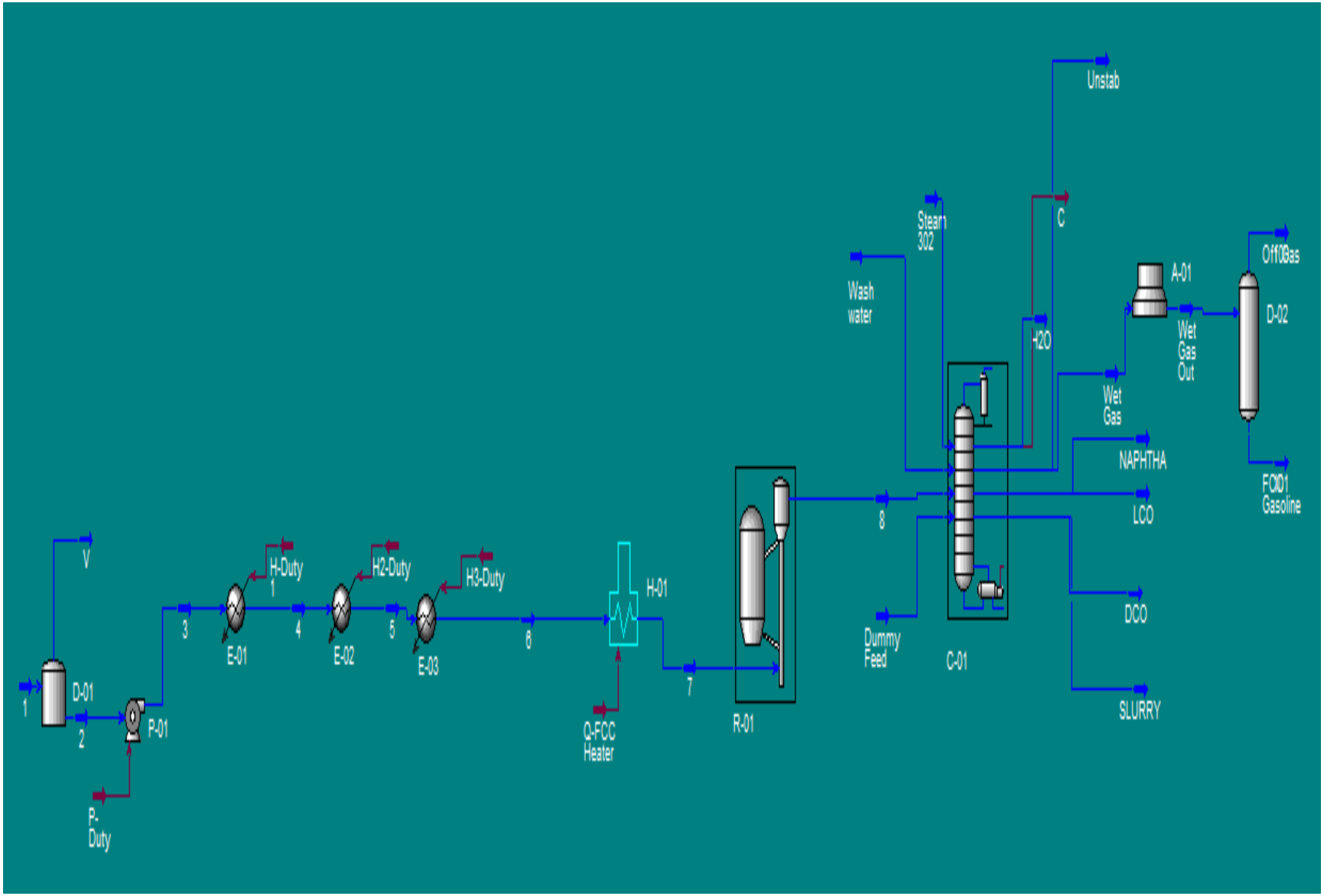


Figure 3.14 Hysys Modeled Pilot FCCU

3.7 Detailed Equipment Design and Specification

Detailed equipment design was carried out with both manual (Mathcad) and Hysys simulation software. The summary is presented in Tables 4.23 to 4.33(result section) while detailed calculations with empherical correlations and design equations used in the manual calculations are presented in appendix C.

3.8 Working Drawings of Individual Equipment

After a detailed design of individual equipment was done, sizing parameters were used to produce working drawing for each equipment for fabrication using computer drawing softwares (AchiCAD and CREO Parametric tool) as shown in figures 4.1 – 4.6. The layout of the plant was also drown.

3.9 Fabrication

Designed equipments were fabricated using standard materials as specified in the equipment specification section. Equipments fabricated are:

1. Fresh Feed Surge Drum (D-01)
2. Fresh Feed/LCO Heat Exchanger (E-01)
3. Fresh Feed/DCO Heat Exchanger (E-02)
4. Fresh Feed/Slurry Heat Exchanger (E-03)
5. Feed Preheater (F-01)
6. FCC Reactor (R-01)
7. Main Fractionator (C-01)
8. Overhead Condenser (A-01)

All materials were sourced locally and fabrication was carried out by a Certified Professional Fabricator in Kaduna under close supervision of both the student and the supervisors. The fabrication stages for each of the equipment is illustrated pictorially in Plates IV - X

However, some design values were not met during the fabrication work as a result of raw material availability and choice for example it is better to use a seamless pipe that match the diameter of a column than to roll a sheet of metal plate therefore, in matching the exact diameter little tolerance is accomadated. Design and actual values used for fabrication are presented in Table 4.34 in the result section.

3.10 Development of Controls, Safety Considerations, Start-up and Shut Down Procedure

In an effort to safe guard the pilot unit, all appropriate control points were identified and control mechanisms were developed to control process parameters like pressure, temperature, flow and level. Essential equipments like the Fired Heater were protected for the safety of both the personnel and the equipment. The start up and shut down procedures were rightly developed bearing in mind the safety of both personnel and the equipment.

CHAPTER FOUR

RESULTS AND DISCUSSION

A pilot plant with capacity to process 5bbl/batch (0.56barrels/hr) with a batch time of 9hrs of Escravos FCC VGO was designed and fabricated. The design followed standard sequence of design procedure using both manual and Hysys simulation software so as to ascertain the accuracy of the manual calculation. The manual calculation was numerically simplified by the use of Mathcad numerical aid.

4.1 Material and Energy Balance

Material and energy balance calculations were carried out using a numerical tool and the result was presented in table Tables 4.1 – 4.11 and both manual and simulation results were presented in Tables 4.12 – 4.22 while a comparism between the manual and the simulation results carried out are shown in Tables 4.12 and 4.13 details are shown in appendices A,B,C and D

Table 4.1: Summary of Material Balance across FCC Unit Surge Drum (D01)

S/N	Components	Input (kg/batch)	Output (kg/batch)
		Stream 1	Stream 2
1	LVGO	112.027	112.027
2	HVGO	100.65	100.65
3	W/Distillate	50.023	50.023
4	C2	0	0
5	C3	0	0
6	C4	0	0
7	Gasoline	0	0
8	LCO	0	0
9	HCO	0	0
10	Slurry	0	0
11	Catalyst	0	0
	TOTAL	262.7	262.7

Table 4.2: Summary of Material Balance across FCC Unit Surge Pump (P01)

S/N	Components	Input (kg/batch)	Output (kg/batch)
		Stream 2	Stream 3
1	LVGO	112.027	112.027
2	HVGO	100.65	100.65
3	W/Distillate	50.023	50.023
4	C2	0	0
5	C3	0	0
6	C4	0	0
7	Gasoline	0	0
8	LCO	0	0
9	HCO	0	0
10	Slurry	0	0
11	Catalyst	0	0
	TOTAL	262.7	262.7

Table 4.3: Summary of Material Balance across FCC Unit Pre-Heater (E01)

S/N	Components	Input (kg/batch)		Output (kg/batch)	
		Stream 3	Stream 11	Stream 4	Stream 17
1	LVGO	112.027	0	112.027	0
2	HVGO	100.65	0	100.65	0
3	W/Distillate	50.023	0	50.023	0
4	C2	0	0	0	0
5	C3	0	0	0	0
6	C4	0	0	0	0
7	Gasoline	0	0	0	0
8	LCO	0	39.405	0	39.405
9	HCO	0	0	0	0
10	Slurry	0	0	0	0
11	Catalyst	0	0	0	0
	TOTAL	302.105		302.105	

Table 4.4: Summary of Material Balance across FCC Unit Pre-Heater (E02)

S/N	Components	Input (kg/batch)		Output (kg/batch)	
		Stream 4	Stream 12	Stream 5	Stream 18
1	LVGO	112.027		112.027	0
2	HVGO	100.65		100.65	0
3	W/Distillate	50.023		50.023	0
4	C2	0		0	0
5	C3	0		0	0
6	C4	0		0	0
7	Gasoline	0		0	0
8	LCO	0		0	0
9	HCO	0	15.762	0	15.762
10	Slurry	0		0	0
11	Catalyst	0		0	0
	TOTAL	278.462		278.462	

Table 4.5: Summary of Material Balance across FCC Unit Pre-Heater (E03)

S/N	Components	Input (kg/batch)		Output (kg/batch)	
		Stream 5	Stream 13	Stream 6	Stream 19
1	LVGO	112.027	0	112.027	0
2	HVGO	100.65	0	100.65	0
3	W/Distillate	50.023	0	50.023	0
4	C2	0	0	0	0
5	C3	0	0	0	0
6	C4	0	0	0	0
7	Gasoline	0	0	0	0
8	LCO	0	0	0	0
9	HCO	0	0	0	0
10	Slurry	0	15.762	0	15.762
11	Catalyst	0	0	0	0
	TOTAL	278.462		278.462	

Table 4.6: Summary of Material Balance across FCC Unit Fired heater (F01)

S/N	Components	Input (kg/batch)	Output (kg/batch)
		Stream 6	Stream 7
1	LVGO	112.027	112.027
2	HVGO	100.65	100.65
3	W/Distillate	50.023	50.023
4	C2	0	0
5	C3	0	0
6	C4	0	0
7	Gasoline	0	0
8	LCO	0	0
9	HCO	0	0
10	Slurry	0	0
11	Catalyst	0	0
	TOTAL	262.7	262.7

Table 4.7: Summary of Material Balance across FCC Unit Reactor (R01)

S/N	Components	Input (kg/batch)		Output (kg/batch)	
		Stream 7	Stream 8	Stream 9	Catalyst
1	LVGO	112.027		0	0
2	HVGO	100.65		0	0
3	W/Distillate	50.023		0	0
4	C2	0		10.508	0
5	C3	0		15.762	0
6	C4	0		31.524	0
7	Gasoline	0		133.977	0
8	LCO	0		39.405	0
9	HCO	0		15.762	0
10	Slurry	0		15.762	0
11	Catalyst	0	37.566	0	37.566
	TOTAL	300.266		300.266	

Table 4.8: Summary of Material Balance across FCC Unit Main Fractionator (C01)

S/N	Components	Input (kg/batch)	Output (kg/batch)			
		Stream 9	Stream 10	Stream 11	Stream 12	Stream 13
1	LVGO	0	0	0	0	0
2	HVGO	0	0	0	0	0
3	W/Distillate	0	0	0	0	0
4	C2	10.508	10.508	0	0	0
5	C3	15.762	15.762	0	0	0
6	C4	31.524	31.524	0	0	0
7	Gasoline	133.977	133.977	0	0	0
8	LCO	39.405	0	39.405	0	0
9	HCO	15.762	0	0	15.762	0
10	Slurry	15.762	0	0	0	15.762
11	Catalyst	0	0	0	0	0
	TOTAL	262.7	262.7			

Table 4.9: Summary of Material Balance across FCC Unit Air Fin Cooler (A01)

S/N	Components	Input (kg/batch)	Output (kg/batch)
		Stream 10	Stream 14
1	LVGO	0	0
2	HVGO	0	0
3	W/Distillate	0	0
4	C2	10.508	10.508
5	C3	15.762	15.762
6	C4	31.524	31.524
7	Gasoline	133.977	133.977
8	LCO	0	0
9	HCO	0	0
10	Slurry	0	0
11	Catalyst	0	0
	TOTAL	191.771	191.771

Table 4.10: Summary of Material Balance across FCC Unit Overhead Separator (D02)

S/N	Components	Input (kg/batch)	Output (kg/batch)	
		Stream 14	Stream 15	Stream 16
1	LVGO	0	0	0
2	HVGO	0	0	0
3	W/Distillate	0	0	0
4	C2	10.508	10.508	0
5	C3	15.762	15.762	0
6	C4	31.524	31.524	0
7	Gasoline	133.977	0	133.977
8	LCO	0	0	0
9	HCO	0	0	0
10	Slurry	0	0	0
11	Catalyst	0	0	0
	TOTAL	191.771	191.771	

Table 4.11: Summary of Energy Balance (Heat Loads) Across All Equipments

S/N	Equipment	Inlet Enthalpy ΔHin (KJ/batch)	Outlet Enthalpy ΔHout (KJ/batch)	Heat Load ΔHout − ΔHin (KJ/batch)
1	Feed Surge Drum (D01)	77.181	38.591	−38.591
2	Feed Pump (P01)	38.591	77.181	38.591
3	Heat Exchanger (E01)	77.181	5016.76	4939.609
4	Heat Exchanger (E02)	5016.79	10265	5248
5	Heat Exchanger (E03)	10270	18909	8644
6	Feed Preheater (F01)	18910	22770	3859
7	Reactor (R01)	22781.172	26510	3732
8	Main Fractionator (C01)	26510	4893	−21620
9	Air Fin Cooler (A01)	1474	246.2	−1228
10	Overhead Separator (D02)	246.2	235.6	−10.61

4.2 Manual and Simulation Results

Table 4.12: Reactor Product Distribution from Aspen Hysys and Manual Design

	Aspen Hysys Weight (%)	Manual (MathCad) Weight (%)	Difference
H_2S	-	-	-
C2- minus	3.22	4	0.78
nC3/iC3	6.1	6	0.1
nC4/iC4	10.06	12	1.94
Naphtha C5- 430F	47.23	51	3.77
LCO 430-650F	16.28	15	1.28
Decant Oil 650F+	11.10	6	4.1
Coke Yield	6.08	6	0.08
Total	100	100	-

As can be seen in Table 4.12 a little disparity of 0.1 was observed in iC3/nC3 yield, others are coke yield, LCO, nC4/iC4 with a difference of 0.08, 1.28 and 1.94 respectively. Also obvious in Table 4.13 is a small variation across values with only a high difference in column plate design in which the manual gives 10 plates while the simulation gives 18 plates resulting to a difference of 8, others are column height, hole diameter, weir length, down comer area and column wall thickness with a difference of 0.625, 0.0025, 0.098, 0.00328 and 0.0038 respectively . All other specification were either same or with a little difference both in product distribution and column design values. This shows that the manual calculation is valid even though different correlations were used (Oche 2015) . This also ascertain the correlations validity used in the manual technique.

Table 4.13: Main Fractionator Aspen Hysys and Manual Design

	Parameter	Aspen Hysys	Manual (MathCad)	Difference
S/N	Tray type	sieve	Sieve	-
1	Number of pass	1	1	0
2	Number of spacer	1	1	0
3	Spacer diameter, m	0.005	0.004	0.001
4	Number of trays	18	10	8
5	Number of holes/tray	134	104	30
6	Column diameter, m	0.52	0.44	0.08
7	Column Height, m	2.925	2.2	0.625
7	Hole diameter, m	0.002	0.0045	0.0025
8	Weir height, m	0.0381	0.05	0.0119
9	Weir length, m	0.054	0.152	0.098
10	Tray spacing, m	0.15	0.2	0.05
11	Plate deflection, mm	< 3	< 3	0
12	Down comer area, m^2	0.000438	0.003717	0.00328
13	Material of construction	Stainless Steel	Stainless Steel	-
14	Column wall thickness, m	0.0012	0.005	0.0038
15	Insulation thickness, m	0.075	0.075	-
16	Insulation material	Mineral Wool	Asbestos	-

From table 4.12 which shows the product distribution, it can be seen that computer simulation result showed a better result when it come to the desired output i.e gasoline with 51% compared

with 47.23% for manual. It was also obvious that less of LCO, DCO and the slurry of 15, 6 and 6 % were obtained compared to 16.28,11.18 and 6.06 % respectively. Computer simulation also gave more of the off gases i.e C_1-C_4 when compared to manual which made the adoption of the simulation values for fabrication.

Table 4.14: Aspen Hysys Reactor Product Distribution

Components	Volume Flow (m^3/hr)	Mass Flow (kg/hr)	Volume (%)	Weight (%)
H_2S	-	-	-	-
Fuel Gas	6.69	5682.9	3.50	3.22
Propane	6.09	3093.3	3.19	1.75
Propylene	14.8	7685.1	7.75	4.35
nButane	3.96	2313.2	2.07	1.31
iButane	9.25	5214.8	4.84	2.95
Butenes	16.8	10242.8	8.80	5.80
Naphtha C5-430F	108.9	83400.3	57.06	47.23
LCO 430-650F	32.38	28737.6	16.95	16.28
Bottoms 650F+	18.07	19597.3	9.46	11.10
Coke Yield		10736.6		6.08
Total	210.4	176571.0	110.11	100.00
Conversion			73.59	72.63

Table 4.15 shows Reactor feed condition for Aspen Hysys which is very important for FCC design and operation. Properties like the CCR controls the reaction and the heat balance within the regenerator and as such it has to be maintained within a range of values depending on feed type (Leuenberger, 1988).

Table 4.15: Aspen Hysys Reactor Feed Condition

Bulk Properties	FCC Feed
API Gravity	21.27
Specific Gravity 60F/60F	0.9262
Basic Nitrogen Content (ppm wt)	461.4
Total Nitrogen Content (ppm wt)	1384
Total/Basic Nitrogen Ratio	3.00
Sulphur Content (%)	1.15
Fraction of Feed Processed	0.50
Conradson Carbon Residue (%)	0.57

Table 4.16 present the reactor geometry showing dimensions of the reactor (i.e the riser, disengager, stripper and regenerator). In this design attension was given to the reactor as it is the heart of the unit. The Catalytic Cracking reactions are sequential processes in which many desirable products are intermediates (Ye-Mon, 2004), as such cracking reactions must be terminated after a desirable reaction time; otherwise, desirable products will continue to crack, leading to excess productions of light gases and coke. In this design better riser termination technology which sharpens the termination of reactions by the combination of the unique design

of primary stripper cyclones and close coupled secondary cyclones (Letzsch et al, 2009) were employed. Moreover, the riser design also takes care of the volume increase as a result of change of phase from liquid to vapour. Others are stripper and regenerator separated by a plug valve which were all taken care of in the design.

Table 4.16: Aspen Hysys Reactor Geometry

Riser	Dimensions
Total Length(m)	0.987
Diameter(m)	0.432
Riser Termination Zone	
Length(m)	0.845
Diameter(m)	0.545
Catalyst Stripper	
Height(m)	0.450
Diameter	0.250
Annulus Diameter(m)	0.154
Regenerator	
Dense Bed Height(m)	0.572
Dense Bed Diameter(m)	1.543
Dilute Phase Diameter	0.343
Interface Diameter(m)	0.320
Cyclone Inlet Height(m)	0.240
Cyclone Inlet Diameter(m)	0.280
Cyclone Outlet Diameter	0.260

Table 4.17 shows simulation result for the main Fractionator with a total number of equilibrium stages of 18 and a feed tray of stage 16 from the top. A top and bottom pump around heat duty of $2.743 * 10^7$ and $1.799 * 10^7$ respectively. Table 4.18 shows Fired Heater result while table 4.19, 4.20, 4.21 and 4.22 shows side strippers, condenser/heat exchanger, surge pump and surge drum respectively.

Table 4.17: Aspen Hysys Simulation Design Parameters for Fractionator

Parameter	Aspen Hysys Simulation Result
No of trays	18
Tray spacing	0.15
Column diameter, m	0.52
Feed stage	Tray 16
LCO draw off stage	8
HCO draw off stage	13
Top pump around draw stage	8
Top pump around return stage	13
TPA heat duty, kJ/hr	2.743×10^7
Bottom pump around draw stage	18
Bottom pump around return stage	17
BPA heat duty, kJ/hr	1.799×10^7

Table 4.18: Summary of Furnace Design Parameters

S/N	Parameter	Feed Heater
1	Heater type	Cylindrical pipe still heater
2	No of tubes in convection section	-
3	No of rows in convection section	-
4	No of tubes in radiation section	-
5	No of Zones	2
6	No of rows in radiation section	-
7	Diameter of tube, m	0.024
8	Length of tube, m	1
9	Spacing between tubes	2D
10	Heater Volume, m^3	0.197
11	Base Diameter, m^3	0.872
12	Heater Height, m	1.745
13	Material of construction	Mild Steel
14	Insulation thickness, m	0.075
15	Lagging Material	Ceramic

Table 4.19: Summary of Column Design Parameters

S/N	Parameter	Stripper 1 & 2	Main Fractionator
1	Tray type	sieve	sieve
2	Number of pass	1	1
3	Number of spacer	1	1
4	Spacer diameter, m	0.005	0.005
5	Number of trays	4	18
6	Number of holes/tray	10	134
7	Column diameter, m	0.21	0.54
8	Hole diameter, m	0.002	0.002
9	Weir height, m	0.0381	0.0381
10	Weir length, m	0.013	0.054
11	Tray spacing, m	0.15	0.15
12	Plate deflection, mm	< 3	< 3
13	Down comer area, m^2	0.0000314	0.000438
14	Material of construction	Stainless steel	Stainless steel
15	Column wall thickness, m	0.0012	0.0012
16	Column cover thickness, m	0.0015	0.0025
17	Insulation thickness, m	0.075	0.075
18	Insulation Material	Mineral Wool	Mineral Wool

Table 4.20: Summary of Hysys Simulated Condenser/Heat exchanger Design Parameters

S/N	Parameter	Condenser (A-01)	Feed-LCO HE (E-01)	Feed- HCO HE (E-02)	Feed Slurry HE (E-03)
1	Type	Fixed-tube plate shell and tube heat exchanger for all four.			
2	Number of tube	3	1	1	1
3	Outer Diameter of tube, m	0.03	0.03	0.03	0.03
4	Length of tube, m	0.33	1	1	1
5	Tube pitch	1.25	-	-	-
6	Tube pattern	Triangular	-	-	-
7	Number of shells	1	1	1	1
8	Bundle diameter, m	0.12	0.03	0.03	0.03
9	Inside Diameter of shell, m	0.26	0.26	0.26	0.26
10	Number of baffle	1	1	1	1
11	Baffle cut, mm	5	3	3	3
12	Average tube-side bulk temperature K	262	311	413.5	349
13	Average shell-side bulk temperature K	488	426	446.5	367
12	Material of construction	Mild Steel for all four			

Table 4.21: Summary of Surge Pump Design Parameters

Pump Head , m	2.372
Power , kW	3.199×10^3
Construction Material	Carbon Steel

Table 4.22: Summary of Surge/Reflux Drums Design Parameters

Parameter	Surge Drum (D-01)	Reflux Drum (D-02)
Type	Horizontal Drum	Horizontal two phase separator
Area m^2	0.197	1.008
Diameter, m	0.631	1.133
Height , m	1.002	0.94
Construction Material	Stainless Steel	Stainless Steel
Mesh Type	Wire Mesh	

4.3 Detailed Equipment Specification

This involve a detailed design of individual equipment manually using MathCad and all sizing parameters were established in this section as shown in Tables 4.23 – 4.33.

Table 4.23: Surge Drum (D-01) Design Parameters

Item Number	001
Equipment	Surge Drum
Equipment Tag	D01
Specification	
Volume	0.197 m^3
Diameter	0.631 m
Height	1.002 m
Construction Material	Stainless Steel

Table 4.24: Surge Pump (P-01) Design Parameters

Item Number	002
Equipment	Surge Pump
Equipment Tag	P01
Specification	
Pump Head	2.372m
Power	3.199 x 103kW
Construction Material	Stainless Steel

Table 4.25: Pre-Heater (E-01) Design Parameters

Item Number	003
Equipment	Heat Exchanger
Equipment Tag	E01
Specification	
Length of Tube	1m
Average tube-side bulk temperature	311 K (38^0C)
Average shell-side bulk temperature	426 K(153^0C)
Construction Material	Stainless Steel

Table 4.26: Pre-Heater (E-02) Design Parameters

Item Number	004
Equipment	Heat Exchanger
Equipment Tag	E02
Specification	
Length of Tube	1m
Average tube-side bulk temperature	413.5 K (140.5^0C)
Average shell-side bulk temperature	446.5 K (173.5^0C)
Construction Material	Stainless Steel

Table 4.27: Pre-Heater (E-03) Design Parameters

Item Number	005
Equipment	Heat Exchanger
Equipment Tag	E03
Specification	
Length of Tube	1m
Average tube-side bulk temperature	462 K (153^0C)
Average shell-side bulk temperature	488 K (215^0C)
Construction Material	Stainless Steel

Table 4.28: Heater (H-01) Design Parameters

Item Number	006
Equipment	Heater
Equipment Tag	H01
Specification	
Volume	0.197 m^3
Diameter	0.872 m
Height	1.745 m
Construction Material	Stainless Steel

Table 4.29: Column (C-01) Design Parameters

Item Number	010
Equipment	Distillation Column
Equipment Tag	C01
Specification	
Tray Spacing	0.2m
Diameter	0.44m
Height	2.2 m
Construction Material	Stainless Steel

Table 4.30: Reactor (R-01) Design Parameters

Item Number	007
Equipment	Reactor
Equipment Tag	R01
Specification	
Volume	0.127 m^3
Diameter	0.545 m
Height	1.817 m
Construction Material	Stainless Steel

Table 4.31. Riser termination dimentsions

Riser	Dimensions
Total Length(m)	0.987
Diameter(m)	0.432
Riser Termination Zone	
Length(m)	0.845
Diameter(m)	0.545
Catalyst Stripper	
Height(m)	0.450
Diameter	0.250
Annulus Diameter(m)	0.154
Regenerator	
Dense Bed Height(m)	0.572
Dense Bed Diameter(m)	1.543
Dilute Phase Diameter	0.343
Interface Diameter(m)	0.320
Cyclone Inlet Height(m)	0.240
Cyclone Inlet Diameter(m)	0.280
Cyclone Outlet Diameter	0.260

Table 4.32: Cooler (A-01) Design Parameters

Item Number	008
Equipment	Condenser
Equipment Tag	A01
Specification	
Length of Tube	1m
Average tube-side bulk temperature	349 K (76^0C)
Average shell-side bulk temperature	367 K (94^0C)
Construction Material	Stainless Steel

Table 4.33: Overhead Separator (D-02) Design Parameters

Item Number	009
Equipment	Overhead Separator
Equipment Tag	D02
Specification	
Area	1.008 m^2
Diameter	1.133m
Height	0.94 m
Construction Material	Stainless Steel

4.4 Working Drawing of Individual Equipment

After a detailed design of individual equipment, data were collected and a working drawing is prepared for each equipment using CREO Parametric Tool for fabrication as shown in figures 4.1 – 4.6. The layout of the plant is also drawn using ArchiCad Tool and all dimensions were rounded up to next available sizes as can be seen in figure 4.7.

4.4.1 Mini FCC surge drum

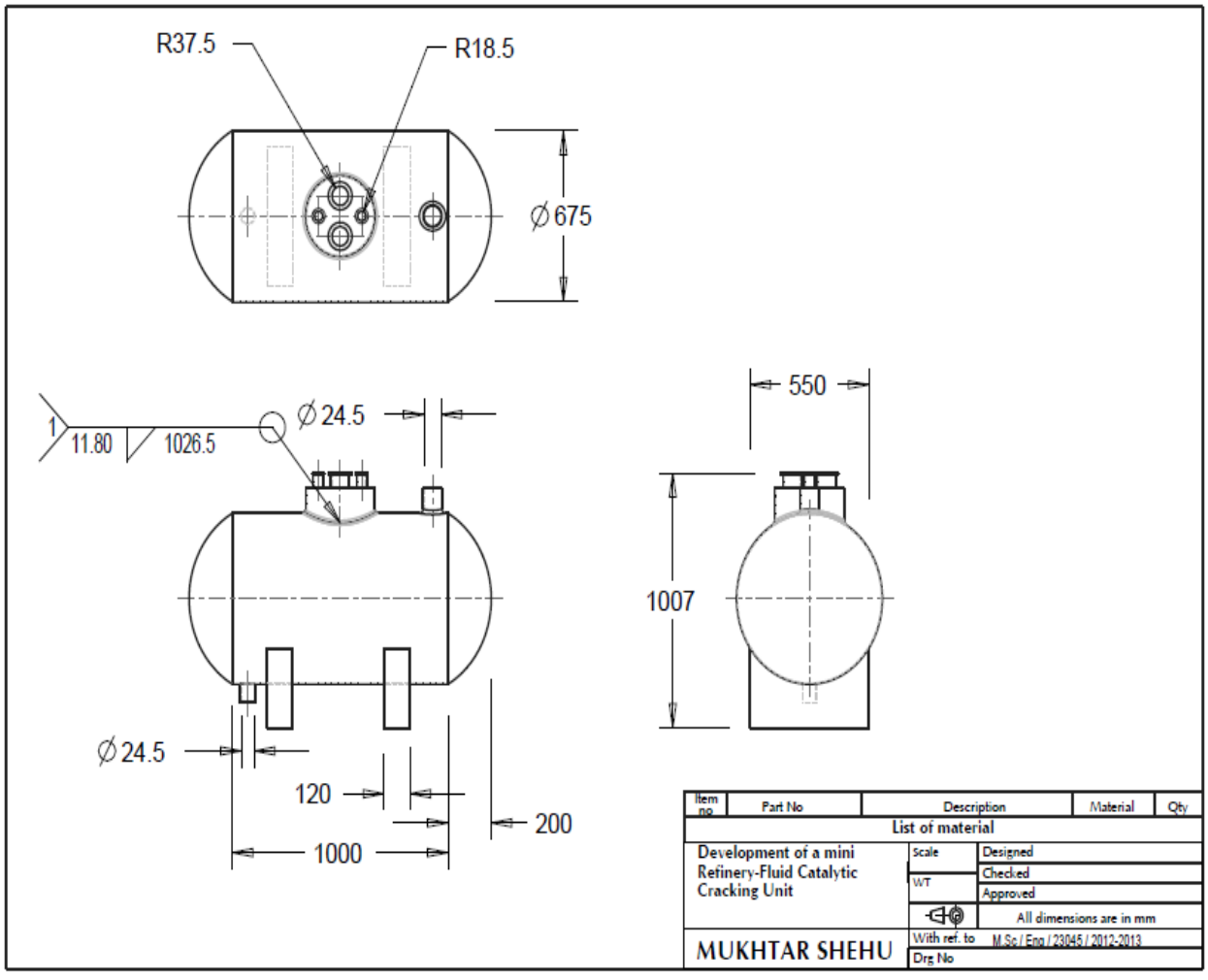


Figure 4.1: Mini FCC Surge Drum

4.4.2 Mini FCC heat exchanger

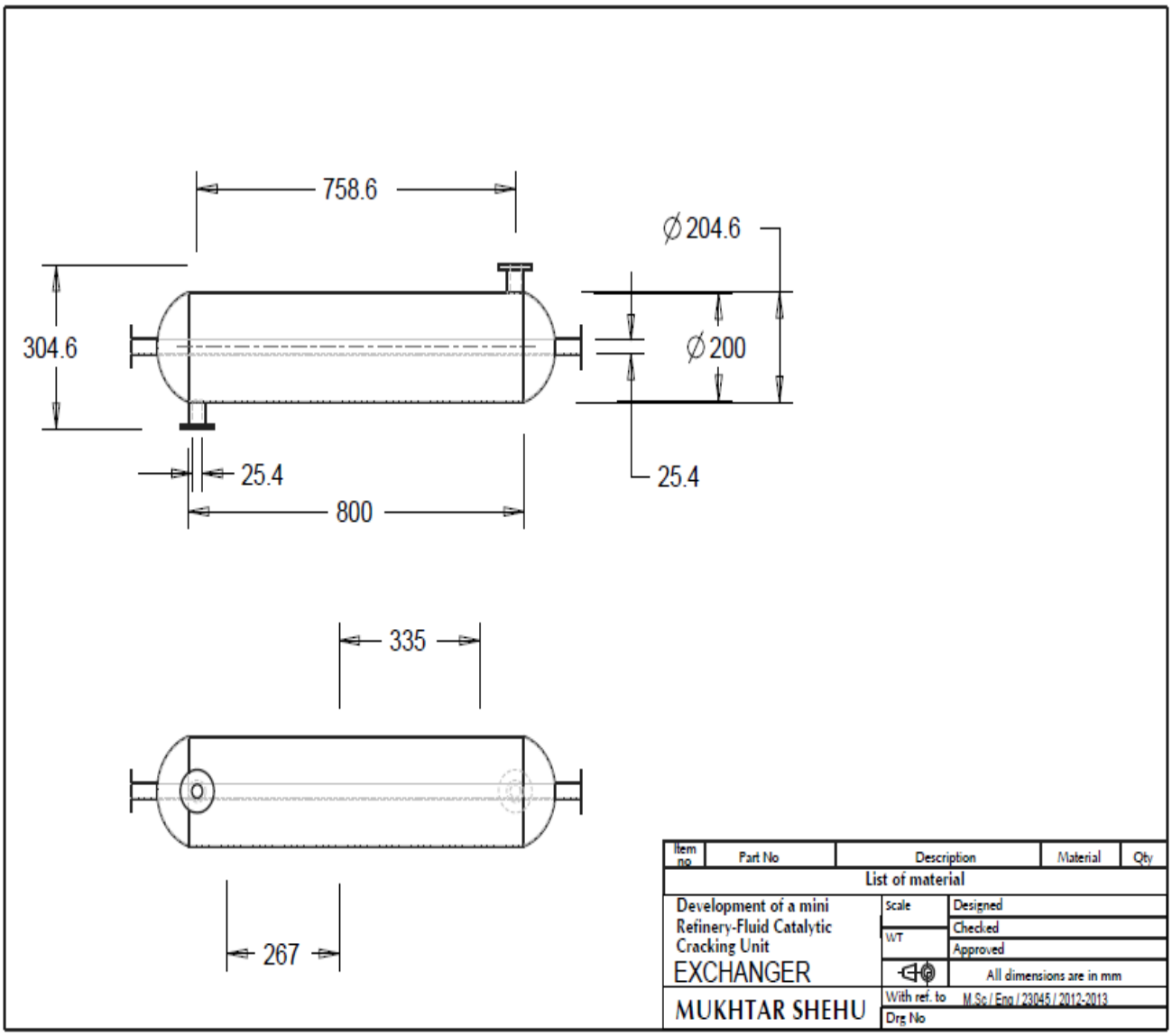


Figure 4.2: Mini FCC Heat Exchanger

4.4.3 Mini FCC fired heater

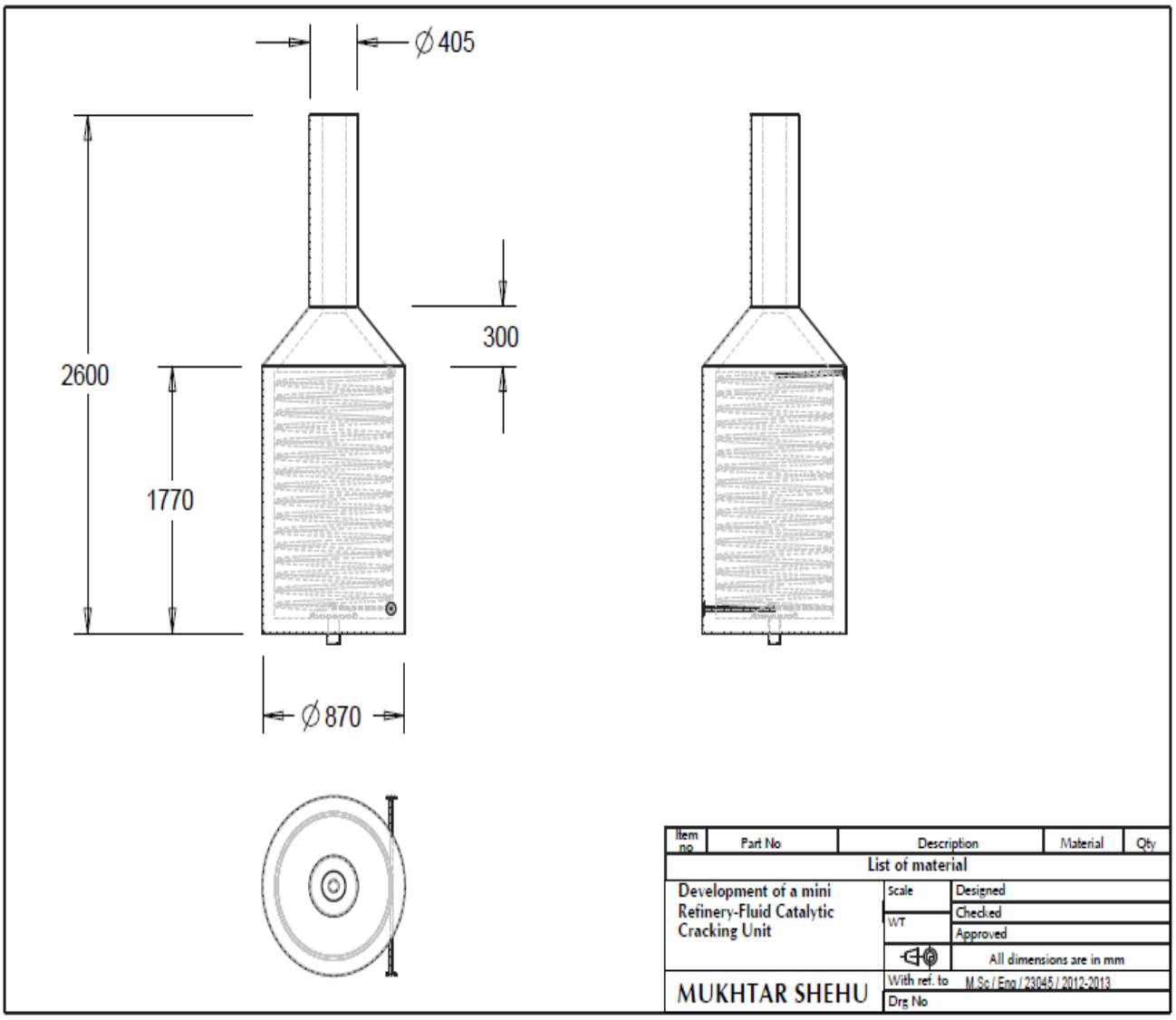


Figure 4.3: Mini FCC Feed Preheater

4.4.4 Mini FCC reactor

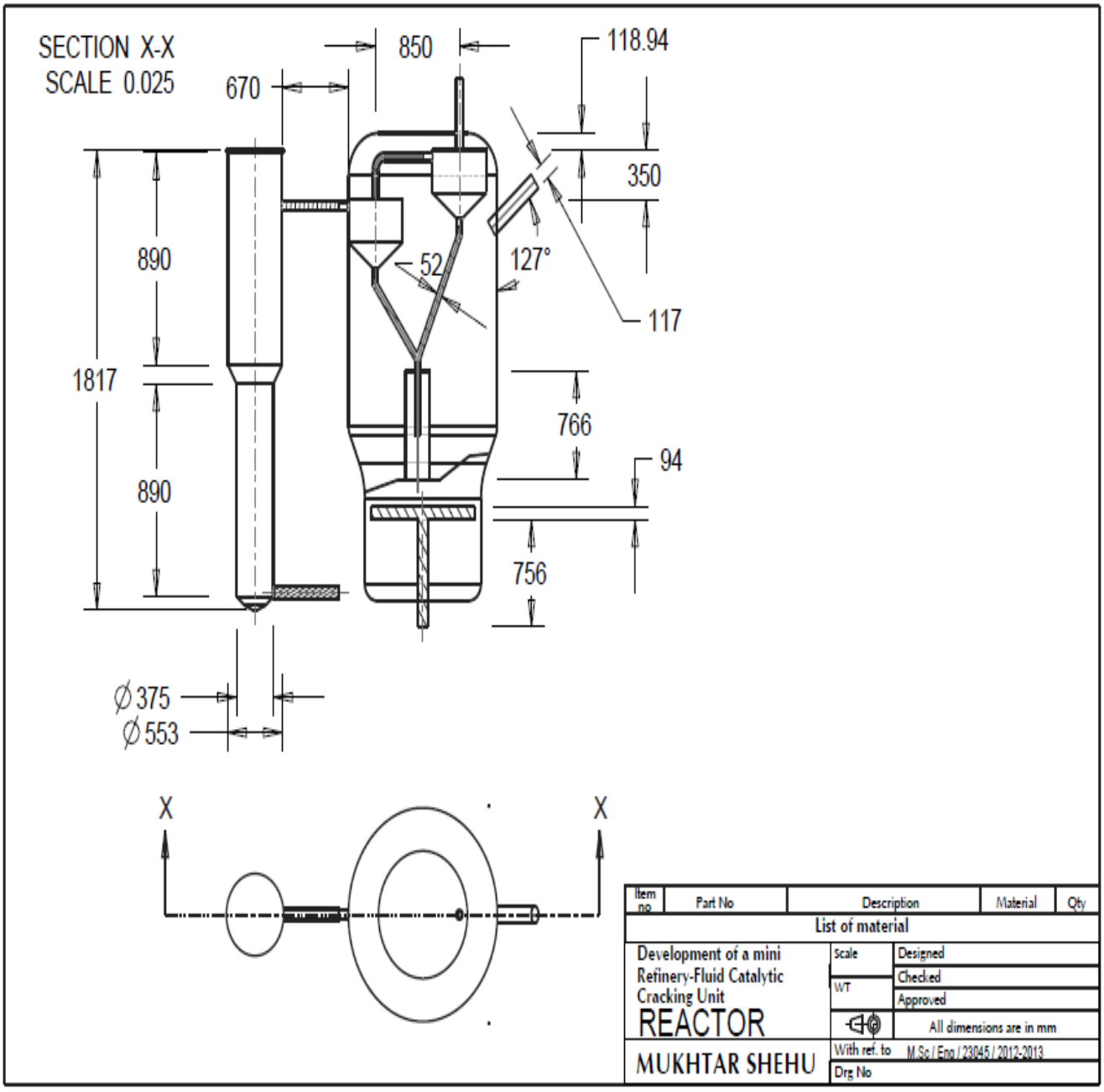


Figure 4.4: Mini FCC Reactor

4.4.5 Mini FCC main fractionator

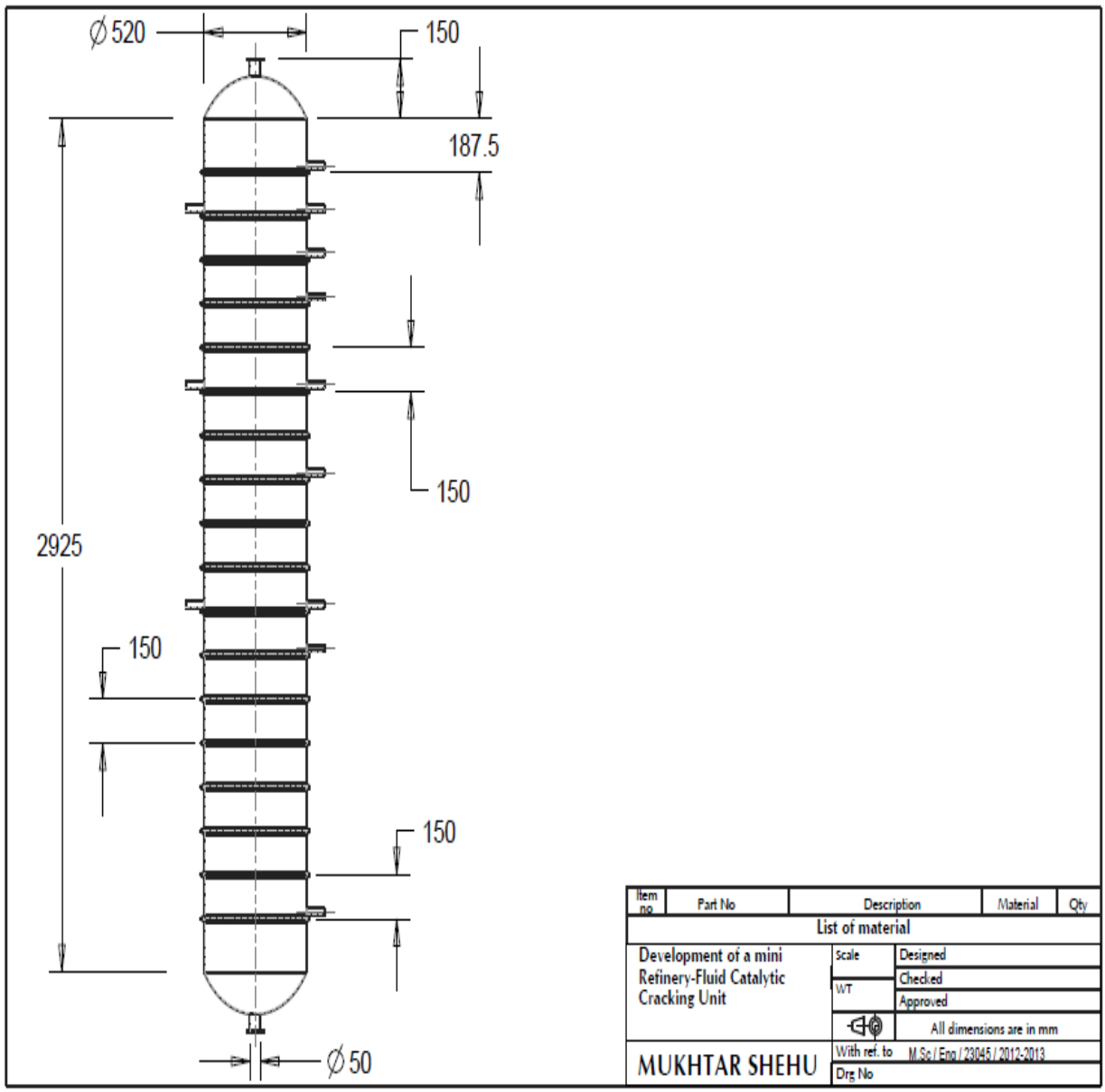


Figure 4.5: Mini FCC Main Fractionator

4.4.6 Mini FCC surge drum

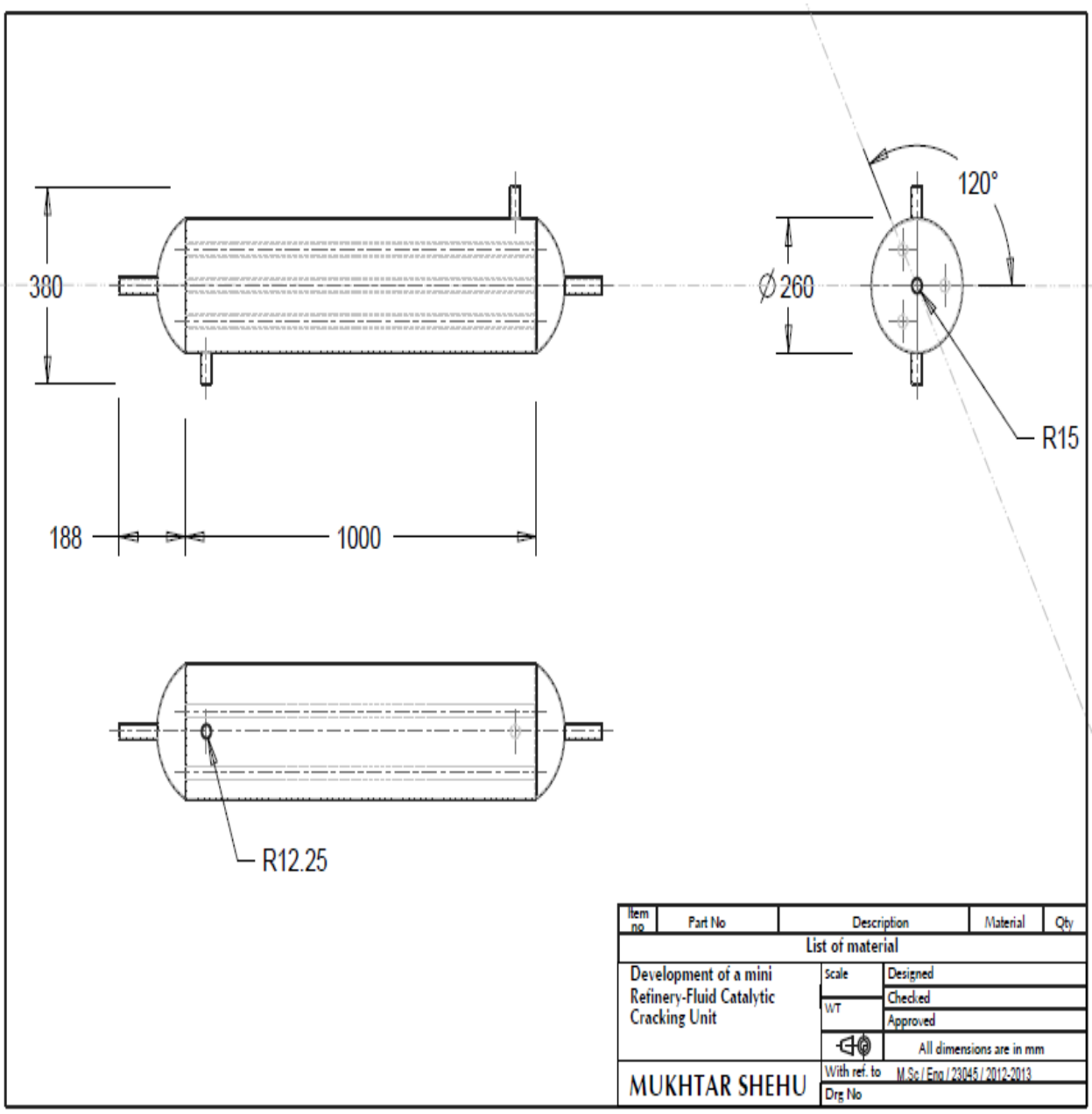


Figure 4.6: Mini FCC Overhead Condenser

4.4.7 Mini FCC plant layout

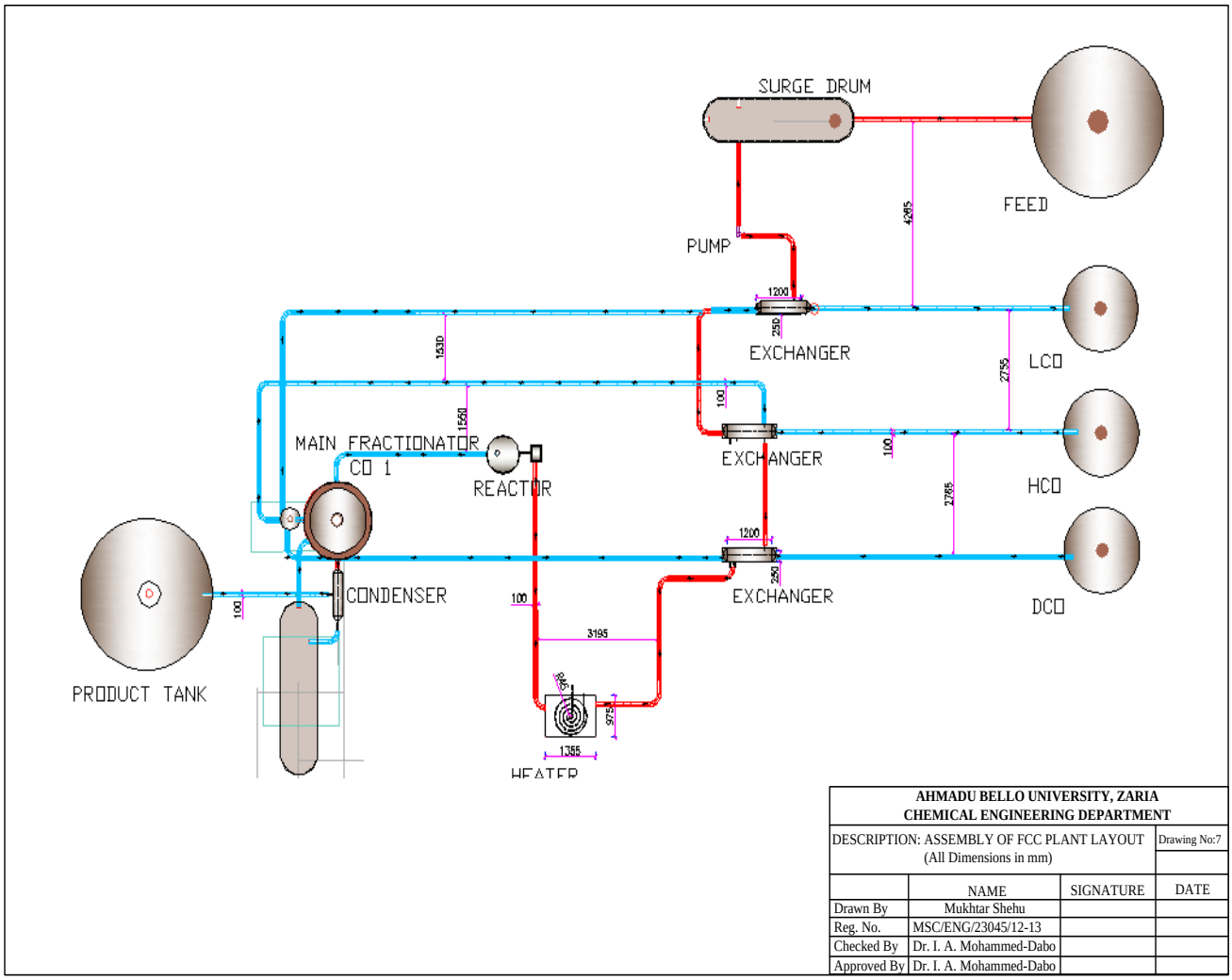


Figure 4.7: Mini FCC Plant Layout

4.5 Fabrication

Fabrication of the designed equipment was carried out in Kaduna by a Certified Professional Fabrication Company. Numerous challenges faced after a lot of rigorosity were solved and the fabrication exercise was successful. Some of the problems encountered were that of material procurement more especially the pipes used in the fabricaton of the heat exchangers and the main fractionator. A seamless pipe with a diameter that matched the designed shell diameter of the

heat exchanger and the thickness has to be searched for and after a wild search, pipes with specifications within the tolerance of the designed parameters were obtained and used for the fabrication as can be seen in table 2.34 and of fabricated equipments plates shown. Other rigorosities were in the aspect of reactor fabrication where the size of the reactor makes the fabrication and installation of the internals difficult.

Table 4.34: Design and actual values used for fabrication

Equipment	**Parameters**	**Design Value**	**Fabricated Value (within tolerance)**
1. Surge Drum	Drum Diameter	0.631m	0.685m
	Drum Height	1.002m	1.200m
	Material of Construction	Stainless Steel	Mild Steel
2. Exchangers	Shell Diameter	0.260m	0.275m
	Tube Diameter	0.030m	0.0245m
	Exchanger Length	1.000m	1.000m
	Material of Construction	Mild Steel	Mild Steel
3. Fired Heater	Base Diameter	0.872m	0.865m
	Height	1.745m	1.562m
	Tube Diameter	0.024m	0.0245m
	Material of Construction	Mild Steel	Mild Steel
4. Main Fractionator	Column Diameter	0.520m	0.510m
	Column Height	2.925m	3.000m
	Material of Construction	Stainless Steel	Mild Steel
5. Overhead Condenser	Shell Diameter	0.260m	0.310m
	Tube Diameter	0.030m	0.0245m
	Condenser Length	1.000m	1.000m
	Material of Construction	Mild Steel	Mild Steel
6. Riser Reactor	Diameter	0.545m	0.540m
	Height	1.817m	1.817

4.5.1 Feed Surge drum



Plate V: Fabricated Surge Drum

4.5.1 Heat exchangers





Plate VI: Fabricated Heat Exchangers with Red Oxide and Aluminium Spray

4.5.3 Overhead condenser





PlateVII: Shell and Tube of Overhead Condenser, before and after Red Oxide Spray

4.5.4 FCC Fired heater



Plate VIII: Fabricated Fired Heater, before and after installation on furnance platform

4.5.5 FCC converter parts



Plate IX: Fabricated Reactor Parts before coupling (a) Stripper (b) 2-Stage Cyclone (c) Riser



Plate X: Fabricated Reactor Parts before coupling (a) Air grid (b) Regenerator (c) Disengager with circmesh

4.5.6 Main fractionator



Plate XI: Fabricated Main Fractionator (a) Sieve tray arrangement (b) Sieve tray (c) Column sprayed with Aluminium Coat (d) Column Sprayed with Red Oxide

4.6 Process Control

Process Control is important in any process plant in which feed material is converted to product through a physical or chemical process. The control variables are basically Temperature, Pressure,Fluid Flow and Level and can be controlled either manually or automatically. Manual control is achieved manually while automatic control involves a sequence of activities from the sensing element (e.g. a thermocouple or an orifice plate) through the transmitter to the main control element (e.g. a control valve).

The control of this pilot unit starts from the Surge drum (D-01) in which a level of not more than 80% is maintained via an automatic control system that connect a sensor in the tank and a control valve at the inlet stream of the drum . Temperature of feed in the feed train exchangers (E-01, E-02 and E-03) can be controlled via a bypass on any of the exchangers and a control valve at the inlet stream of any of the exchanger. Flow of feed into the heater (H-01) also known as heater passes is controlled via a control valve and the heater outlet temperature is controlled by an automatic controller that uses a thermocouple to sense outlet temperature and a valve to regulate the fuel flow into the combustion chamber of the heater.

The reactor is the most complicated section of the unit where a lot of activity is happening. Therefore, all activities need optimum control, these include fluidization aid steams, regeneration air flow, catalyst flow using differential pressure and the plug valve, flue gas flow using the slide valves e.t.c. All steams into the reactor have a temperature indicator while stripping steam into the stripping section of the reactor has a control valve for flow regulation. Spent catalyst movement from the stripper to the regenerator is controlled by a plug valve (PV1) and regenerated catalyst movement from the regenerator section of the reactor to the riser is controlled by a plug valve (PV2) and the entire circulation is made possible by establishing a

pressure differential between the disengager and the regenerator. This is achieved by controlling the flow of flue gas out of the regenerator and the catalyst movement from the disengager through the stripper to the regenerator. Regeneration temperature is controlled by an automatic controller that uses a thermocouple to sense the temperature and a control valve that regulates the air flow into the regenerator.

The Main Fractionator (C-01) is connected to the reactor and therefore a uniform pressure needs to be maintained at the top of the column. Hence, a pressure relief valve (PSV) is needed to maintain a set pressure within the system. Reflux and recycle streams into the column are controlled via remote control valves that are controlled from the control room. A temperature indicator is used to indicate temperature at the inlet and outlet of the overhead condenser (A-01) and a level controller is used to control the level in the overhead separator (D-02) by sensing the level in the drum and opening or closing a control valve on the product discharge line. All product lines (FCC Gasoline, LCO, DCO, and Slurry) will have a temperature indicator prior to storage as can be seen in figure 4.8.

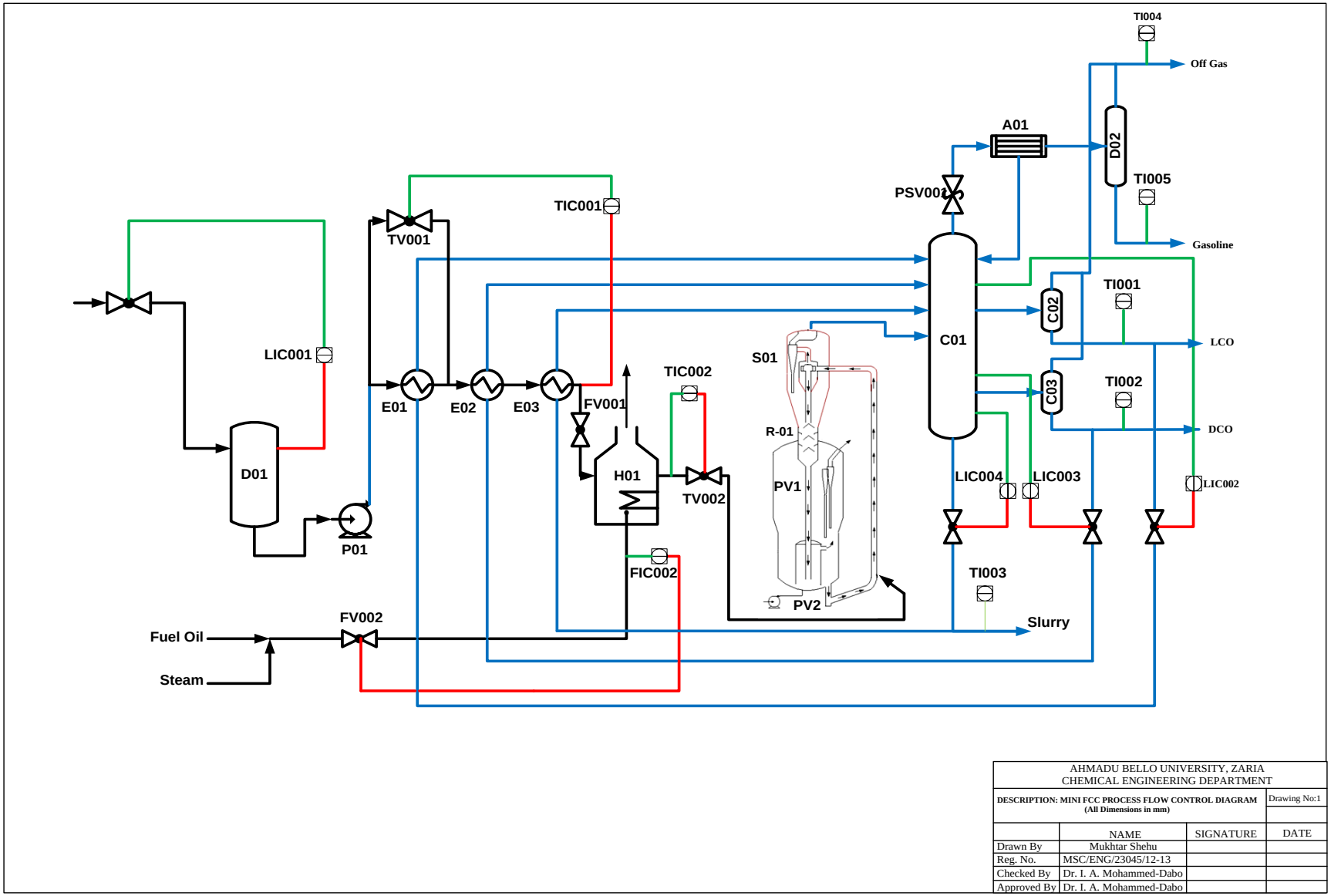


Figure 4.8: FCC Process Flow Control

4.7 Safety Consideration

The safety of a process plant depends solely on the process parameters and how well this parameters are controlled. This parameters are mainly temperature, pressure, level and flow.Safety considerations regarding this design will involve special control systems to be installed to take care of any unsafe condition that may result to accident, explosion or fire. This controls will be installed on some essential equipments that are prone to this hazard. Sections of the plant that will be covered in this work include feed preheating circuit, the converter and the main Fractionator.

4.7.1 Feed preheating circuit

It all start from the feed pump where a control system is employed to monitor the flow through the heater passes and respond by controlling the amount of steam that passes through the turbine thereby increasing flow in the passes (Figure 4.9). This arrangement protect the heater pipes from coking and being heated empty and also avoid shut down in case of either pump failure.

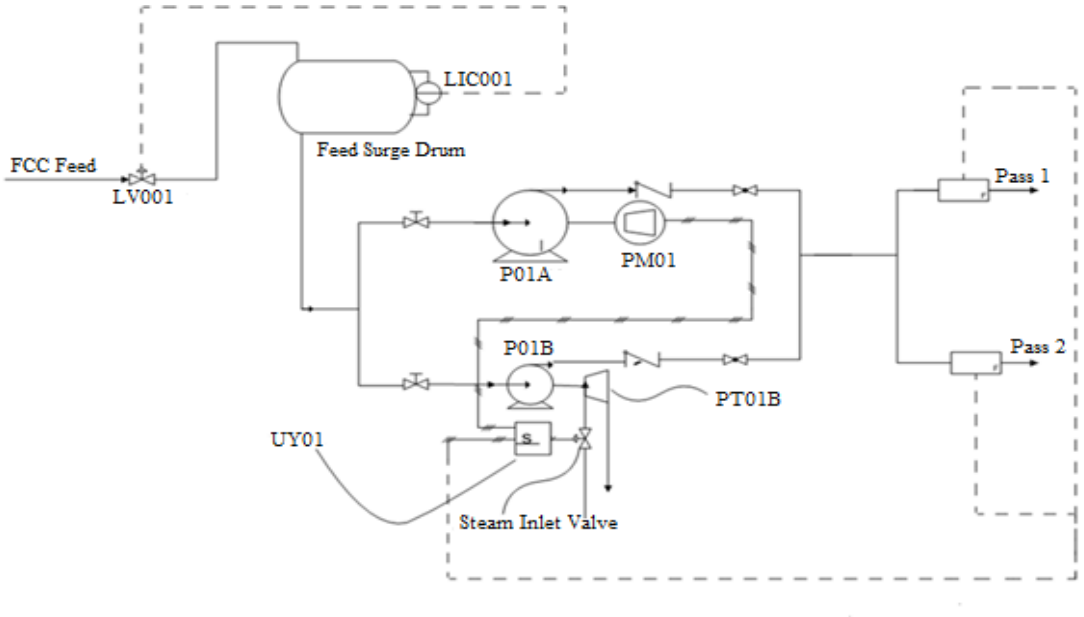


Figure 4.9: Proposed Feed Pump Instrumentation

. The proposed instrumentation diagram for the feed pump is illustrated in Figure 4.9, while figure 4.10 shows that of feed preheater or simply feed fired heater.

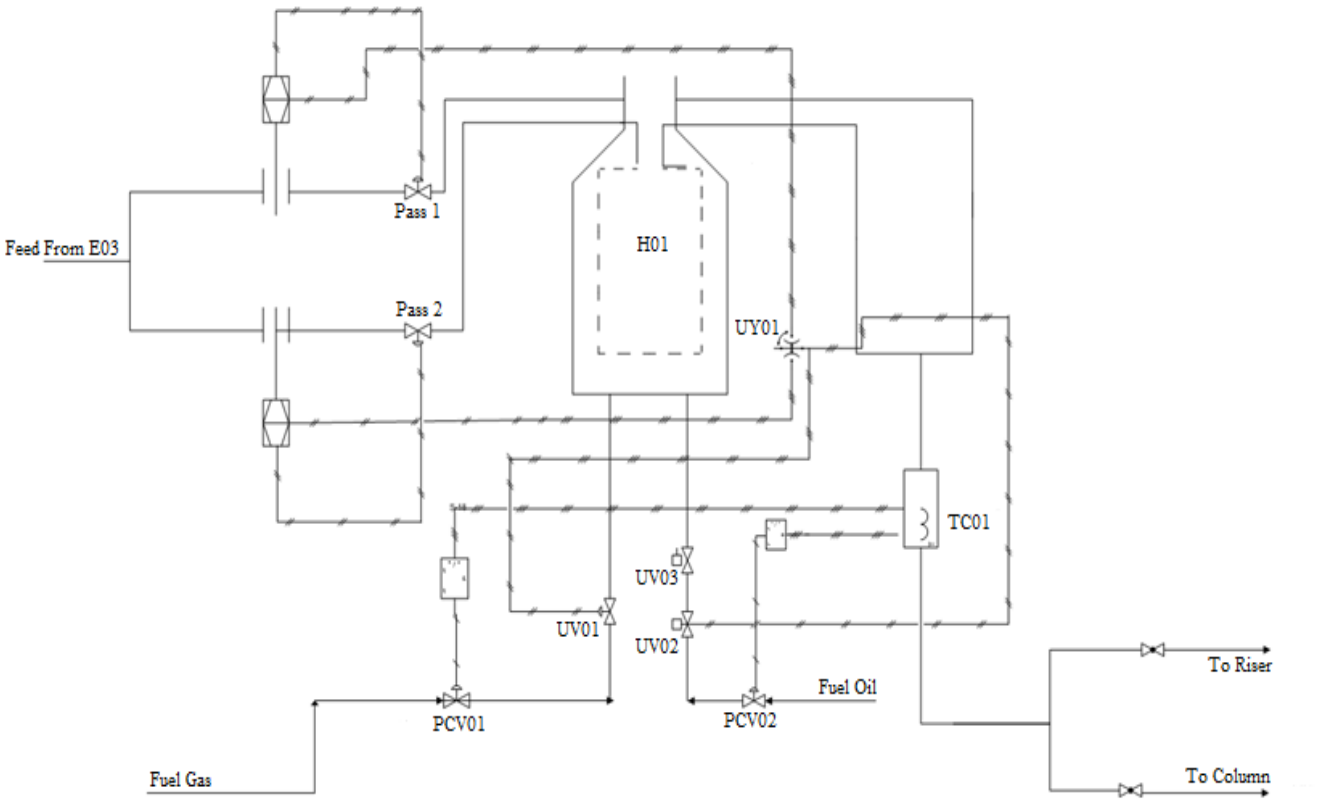
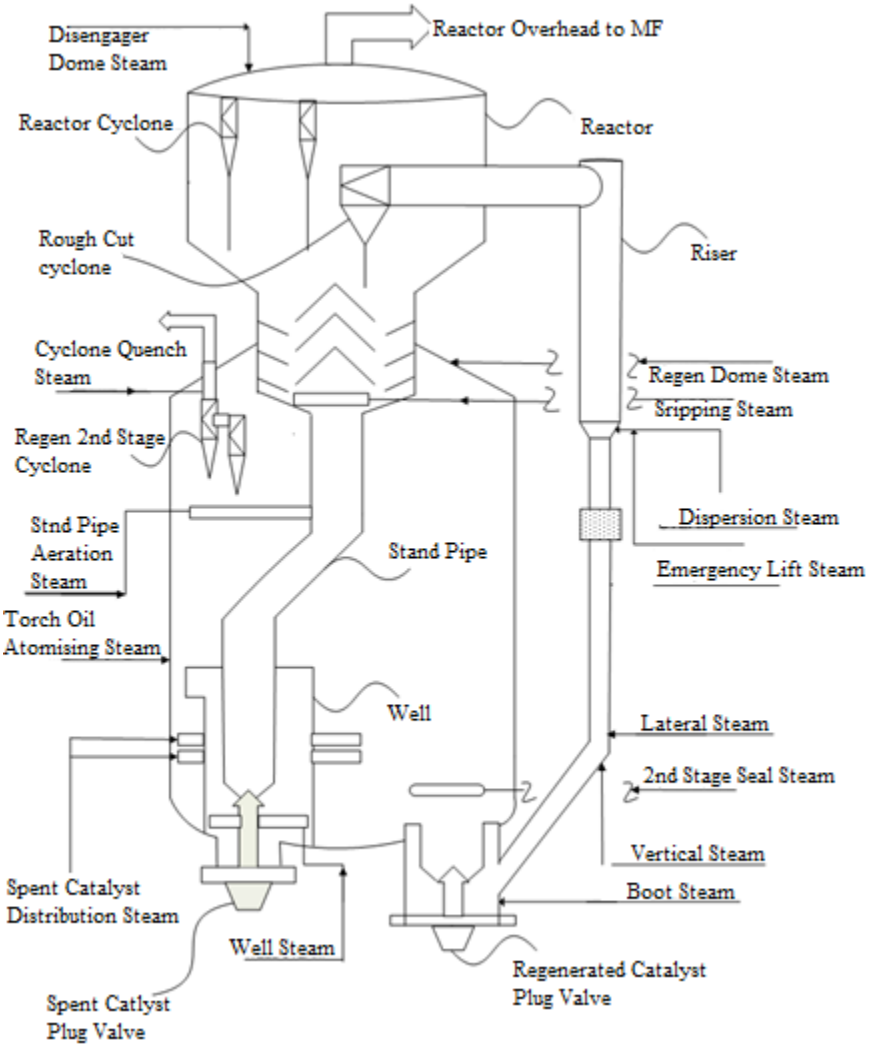


Figure 4.10: Proposed Feed Preheater Instrumentation

The feed preheater is protected by controls as shown in Figure 4.10 where the outlet temperature is sensed and the pressure of both the fuel gas and fuel oil is controlled by PCV01 and PCV02. UY01 which controls the turbine speed of P01 (Figure 4.7.1) is also cascaded to UV01 and UV02 to also control the flow fuel into the heater.

4.7.2 Converter

The Converter is the heart of the unit and its control is important to the safety of the entire unit. Essential component of converter safety include pressure differential control, steam injection and temperature control. Different steam injection point in the converter is shown in Figure 4.11.



Source: (KRPC 1980)

Figure 4.11: Converter Steam Injection Points

All the steam injection points are medium pressure steam at 15kg/cm^2 (15atm) with the exception of the emergency lift steam which is a high pressure steam at 42kg/cm^2 (42atm) used to prevent flow of catalyst back to the air blower in case of blower failure.

Pressure differential is achieved by the use of the two plug valves and the flue gas slide valve to pressurize the regenerator and create a differential between the regenerator and the disengager. A controller ensure the differential by sensing the level of catalyst in the stripper and opening the spent catalyst plug valve to allow flow down to the regenerator and still maintain the differential to avoid flow reversal.

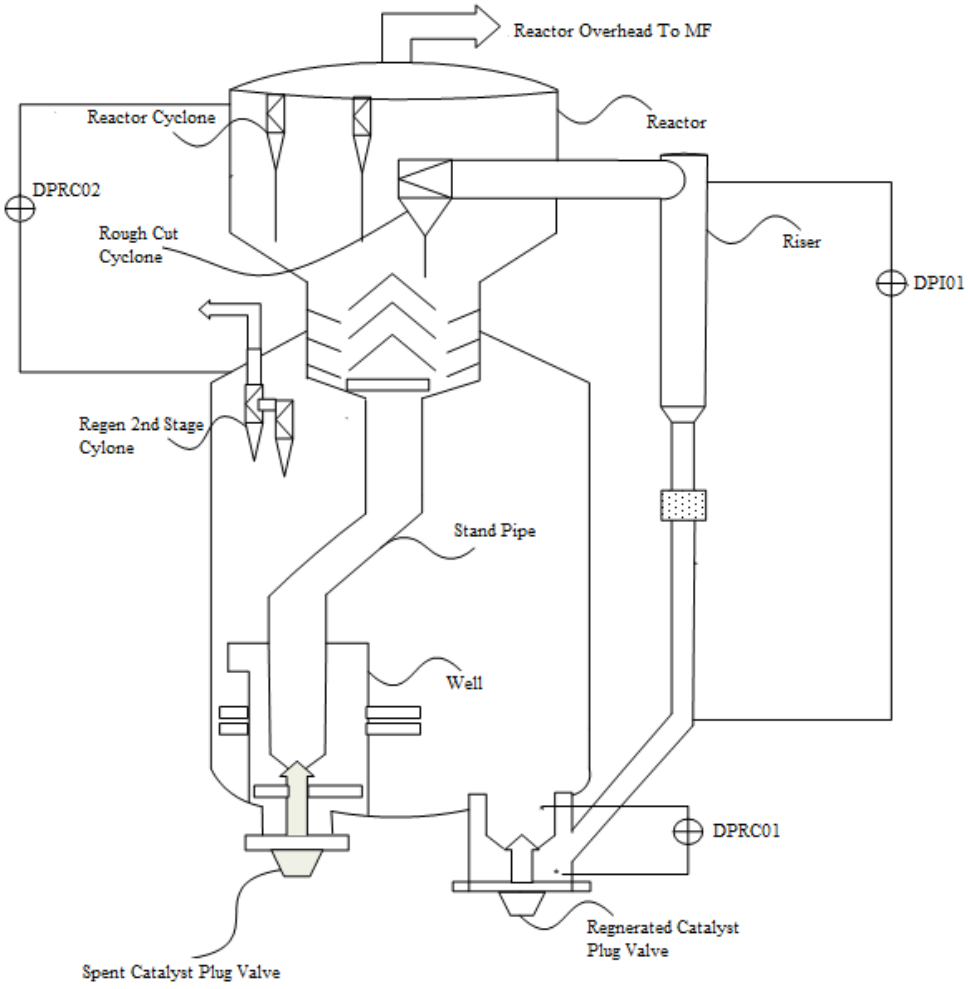


Figure 4.12: Propsed Converter Pressure Tapping Points

Environmental regulation mandates reduction in flue gas particulate emission which will be achieved by the use of an orifice chamber. In addition, the regulation also requires major reduction in flue gas NOx emission, which is typically, achieved by the use of a CO boiler

section and incinerator that burn off the poisonous flue gases to CO_x but in this case the poisonous gas will be bunt off in an incinerator. Conventional technology for flue gas particulate removal includes electro-static precipitators (ESP) and wet scrubbers.

4.7.3 Main fractionator

The control of Column pressure is possible by manipulating the overhead vapour of the column. This is done by bleeding the wet gas to the flare via PCV03 which is controlled by PIC01 as illustrated in Figure 4.13. The Column is protected by pressure safety valve set at 3kg/cm^2 (3atm), any pressure above the set pressure will make the PSV to lift and relieve the pressure thereby protecting the column.

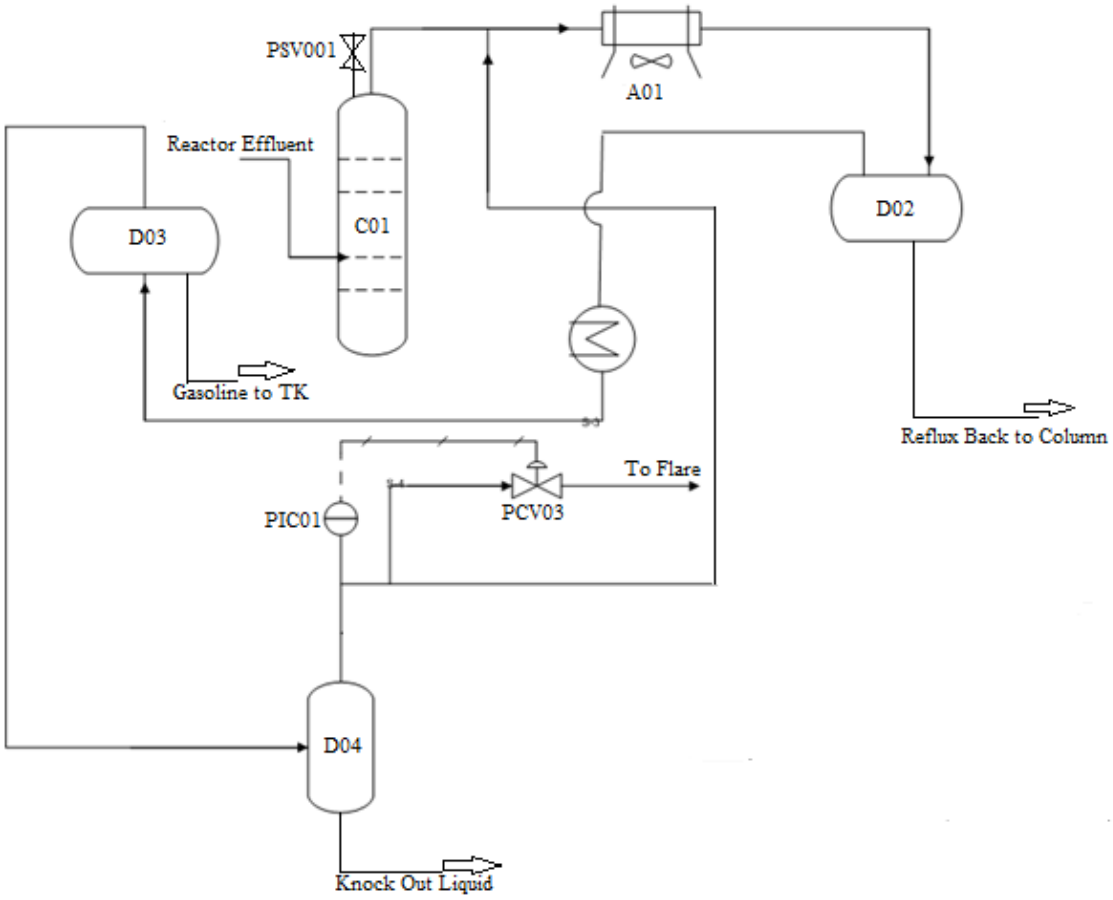


Figure 4.13: Proposed Main Fractionator Pressure Control

The separation of the liquid hydrocarbon is achieved through the use of a reflux drum (D02), the overhead product drum (D03) where Gasoline is sent to tank and the uncondensed gas is further passed through a knock out drum (D04). The knock out liquid is sent back to the overhead Product drum and the wet gas is partly sent to the overhead of the column for pressure control and partly bleeded to flare.

4.8 Start up Procedure

The start up of the plant can be summarized according to the following procedure:

General preparation and confirmations:

This involves availability of all utilities ready for work, all equipments prepared and ready for work and all service units ready for action.

Commissioning of the Air Blower wil follow standard procedure before lighting the air heater.

Start raising the reactor temperature by heating at 50^0C per hr.

Hot bolting at 300^0C and start importing Catalyst into the Reactor

Establishing cold followed by hot circulation at the bottom circuit using necessary bottom pumps

When Disengager outlet temperature is 260^0C, commissioning of medium pressure Steam to the reactor which serve as instrument purge to instrument tappings will be carried out.

When 10% Catalyst level is attained, start catalyst circulation by establishing a pressure differential across the regenerator and the disengager.

When Disengager bed temperature is at least 450^0C and Regenerator bed temperature is at least 540^0C the feed will be commissioned to the Riser to start the process of reaction in the Riser, separation in disengager and stripper and catalyst regeneration in the regenerator.

4.9 Shut Down Procedure

The shut down of the plant can be summarized according to the following procedure:

In the following procedures, it is assumed that the unit is to be shutdown for a general periodic equipment inspection and maintenance. Advance notice should be made so that necessary preparation can be made to permit the unit to be shutdown safely and with minimum loss of production.

It should be noted that, due to the integration of the unit with the other process units, the shutdown operation should be well coordinated with the related departments.

In a normal shutdown, depending on the degree of mechanical work required, it may not be necessary to pump out or steam out all the equipment. In some instances, only one equipment may be involved, and in these cases the necessary equipment will be blanked off in accordance with normal refinery practice in order to minimize the time required for restarting. The unit supervisor should determine which portions of the unit are to be isolated.

Notify supervisory personnel in other departments of the refinery that the FCC unit will be shutdown.

Check that slop and HGO flushing systems are ready for use.

Notify the topping process units personnel that the direct charge to the FCC unit should be gradually stopped.

1. **Reduce fresh feed rate**

In anticipation of shutdown gradually reduce the Fresh Feed rate towards 60% design, via the heater passes. The fresh feed reduction should be carried out slowly enough to adjust operating conditions to maintain product specifications.

As the fresh feed rate is lowered, slowly increase the riser emergency lift steam, proportionately to maintain the riser velocity.

Reduce the Riser Outlet temperature to about 480^0C as the feed rate approaches 60% design.

2. **Adjust regeneration air rate**

Reduce the regeneration air rates to prevent high oxygen content in the flue gas and after-burning as the fresh feed rate is reduced.

3. **Adjust fractionators section**

As the Fresh Feed Rate is reduced, reduce the various fractionators product withdrawals, stripping steams and refluxes

Reduce the Feed Preheat temperature at the outlet of the Heater towards 250^0C. Reduce the burner firing as required ensuring stable firing condition.

Adjust the outlet temperature of the condenser A01 to normal.

Preparation for Cut Out of Fresh Feed with the completion of Fresh Feed reduction toward 60% design, the following actions should then be taken:

Divert overhead receiver product to the light slop header.

Divert LCO product to the Heavy Slop header.

Divert slurry oil to the heavy slop header

Switch flushing oil medium from HCO to Fresh Feed Oil.

Switch seal oil medium from LCO to Fresh Feed Oil.

Switch instruments to Manual Control as necessary.

Corrosion inhibitor injection may be stopped at any convenient time during feed reduction towards zero.

4. **Cut out recycle feed to the Converter**

Stop the recycle feeds to the converter, if being used. Before the fluid in the recycle feed lines congeals, purge out these lines to the converter by connecting a steam hose to convenient drain connections. Then isolate the lines from the Fractionator and Converter, and depressure the lines to the atmosphere.

Continue to lower the Fresh Feed Rate to a minimum via the heater passes while opening feed pump P01 minimum flow bypass. Continue to increase the emergency lift steam and dispersion steam to hold riser velocity. Cut back Heater firing towards shutdown.

Finally stop the feed to the Riser and direct it to the Fractionator, when the fresh feed rate is lowered to 20m^3/hr. Purge out the riser feed line before the line congeals.

Stop firing of the fired heater completely. Shutdown the fired heater according to the operating manual.

4.10 Contribution to Knowledge/ Novelty of the Work

For a Fluid Catalytic Cracking Unit this work has estsblished the effectiveness of Hysys software over manual design through Comparism of design values.

The work also provide fabricated equipments for installation of a pilot FCC unit with the following improvements in the reactor design:

i- Introduction of quench fluid at the riser upper outlet so as to avoid post riser cracking

ii- Development of double side stripper Steam distributor for the prevention of steam back flow.

iii- Introduction of inclined plates (mesh) to replace the Chinese Cap for the prevention of dead zone and an increase in retention time

CHAPTER FIVE

CONCLUSIONS AND RECOMMENDATIONS

5.1 CONCLUSIONS

A mini Fluid Catalytic Cracking Unit with a capacity of processing 5bbl/batch was designed and fabricated. Detailed equipment design using manual and Hysys software was carried out. Process control and safety considerations of the entire unit was also carried out. Although Hysys simulation shows more detailed and effective results, manual design technique gave result close to Hysys simulation. Hence, the manual design technique can be used for FCCU design in the absence of the software.

Equipment fabricated include Feed Surge Drum (Diameter 0.631m and Height 1.002m), Feed/LCO Heat Exchanger (Shell Diameter 0.260m, Tube Diameter 0.030m and Exchanger Length 1.000m), Feed/DCO Heat Exchanger(Shell Diameter 0.275m, Tube Diameter 0.0245m and Exchanger Length 1.000m) , Feed/Slurry Heat Exchanger(Shell Diameter 0.260m, Tube Diameter 0.024m and Exchanger Length 1.000m) , Feed Fired Heater (Base Diameter 0.872m, Height 1.745m and Tube Diameter 0.024m), Riser Reactor (Diameter 0.545m and Height 1.817m), Main Fractionator (Column Diameter 0.520m and Column Height 2.925m) and Overhead Condenser (Shell Diameter 0.310m, Tube Diameter 0.0245m and Condenser Length 1.000m).

The Fabrication was carried out by a certified professional fabrication company using locally sourced materials within the confines of the work.

5.2 RECOMMENDATIONS

The following recommendations should be considered:

i- Installation, piping, instrumentation and test running of the fabricated equipments should be carried out.

ii- The pilot plant when installed should be used for research and development that address salient design problems in Fluid Catalytic Cracking and catalyst enhancement processes.

iii- Other Units such as the Gas Treatment Unit (GTU) and Gas Concentration Unit (GCU) for proper handling of the overhead gas and flue gases should be looked into.

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APPENDIX A

MATERIAL BALANCE

Appendix A entails an illustration of material balance calculations using MathCad numerical tool with an example of the surge drum other equipments follow the same procedure.

Definition: Species indices in the order 1-LVGO, 2-HVGO, 3-$W_{Distillate}$ 4-C_2, 5-C_3, 6-C_4, 7-Gasoline, 8-LCO, 9-DCO, 10-Slurry, 11-Catalyst

Component indices $s := 1..11$ ORIGIN≡ 1

Stream indices $j := 1..16$

Stream initialization:

Mass flow of a component **s** in a given stream **j** is

$$n_{s,j} := 0 \cdot \frac{kg}{batch}$$

Total mass flow of a given stream **j** is

$$N_j := 0 \cdot \frac{kg}{batch}$$

Mass composition of components **s** in stream **j** is

$$x_{s,j} := 0$$

Molar flow of a component **s** in a given stream **j** is

$$z_{s,j} := 0 \cdot \frac{kmol}{batch}$$

Total molar flow of a given stream **j** is

$$Z_j := 0\,\frac{kmol}{batch}$$

Molar composition of components **s** in stream **j** is

$$y_{s,j} := 0$$

$$kJ \equiv 1000\,J \qquad ORIGIN \equiv 1 \qquad kmol \equiv 1000\,mol \qquad \lambda_s := 0 \cdot \frac{kJ}{kmol}$$

Feed Stock (FCC Feed) Compositions:

	$x_{s,1} :=$
LVGO	0.4264
HVGO	0.3831
W Distillate	0.1904
C_2	0
C_3	0
C_4	0
Gasoline	0
LCO	0
DCO	0
Slurry	0
Catalyst	

$$\sum_{s=1}^{11} x_{s,1} = 1$$

Problem Analysis:

Assuming a batch is 9 hours per batch

$$\text{barrel} \equiv \frac{\text{tonne}}{7.312}$$

$$\text{batch} \equiv 9 \cdot \text{hr}$$

From the Given problem statement, volumetric production rate of FCC Unit products is

$$\text{Scale_up_Factor} := 0.38421$$

$$\text{Basis} := \text{Scale_up_Factor} \cdot \left(5 \cdot \frac{\text{barrel}}{\text{batch}} \right)$$

Molecular weight of the components

$$W_s := \begin{pmatrix} \text{LVGO} \\ \text{HVGO} \\ \text{Distillate} \\ C_2 \\ C_3 \\ C_4 \\ \text{Gasoline} \\ \text{LCO} \\ \text{DCO} \\ \text{Slurry} \\ \text{Catalyst} \end{pmatrix} \qquad \text{Molwt} := \begin{pmatrix} 214 \\ 265 \\ 266 \\ 30 \\ 44 \\ 58 \\ 72 \\ 86 \\ 100 \\ 114 \\ 128 \\ 234 \end{pmatrix} \qquad M_{W_s} := \text{Molwt}_s \cdot \frac{\text{kg}}{\text{kmol}}$$

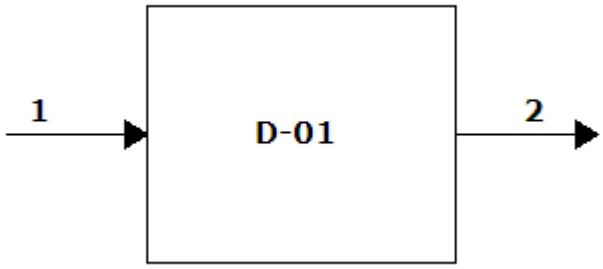
M_{W_S} =

0.214	$\frac{kg}{mol}$
0.265	
0.266	
0.03	
0.044	
0.058	
0.072	
0.086	
0.1	
0.114	
0.128	

FEED STOCK

LVGO	$x_{s,1}$:=	0.4264
HVGO		0.3831
W Distillate		0.1904
C_2		0
C_3		0
C_4		0
Gasoline		0
LCO		0
DCO		0
Slurry		0
Catalyst		

Unit 1: FCC Unit Feed Storage Tank D-01



Stream 1:

Assumption:

1. 100% material recovery

Stream 1 is calculated as:

$n_{s,1} := x_{s,1} \cdot Basis$

$n_{s,1} =$

LVGO	112.027	$\cdot \frac{kg}{batch}$
HVGO	100.65	
W	50.023	
Distillate	0	
C_2	0	
C_3	0	
C_4	0	
Gasoline	0	
LCO	0	
DCO	0	
Slurry	0	
Catalyst		

Input:

$$N_1 := \sum_s n_{s,1}$$

$$N_1 = 262.7 \frac{kg}{batch}$$

Since there is no change of material, stream 2 = stream 1

Stream 2:

$$n_{s,2} := n_{s,1}$$

$$n_{s,2} = \begin{pmatrix} \text{LVGO} \\ \text{HVGO} \\ \text{Distillate} \\ C_2 \\ C_3 \\ C_4 \\ \text{Gasoline} \\ \text{LCO} \\ \text{DCO} \\ \text{Slurry} \\ \text{Catalyst} \end{pmatrix} \begin{pmatrix} 112.027 \\ 100.65 \\ 50.023 \\ 0 \\ 0 \\ 0 \\ 0 \\ 0 \\ 0 \\ 0 \\ 0 \end{pmatrix} \cdot \frac{kg}{batch}$$

Total amount of stream 2 is:

$$N_2 := \sum_s n_{s,2} \qquad N_2 = 262.7 \frac{kg}{batch}$$

Table 4.1: Summary of Material Balance across Feed Surge Drum D-01

Components:	**INPUT**	**OUTPUT**
	Stream 1	**Stream 2**
	$n_{s,1} =$	$n_{s,2} =$
LVGO	112.027 $\cdot \frac{kg}{batch}$	112.027 $\cdot \frac{kg}{batch}$
HVGO	100.65	100.65
W	50.023	50.023
Distillate	0	0
C_2	0	0
C_3	0	0
C_4	0	0
Gasoline	0	0
LCO	0	0
DCO	0	0
Slurry	0	0
Catalyst	0	0
Total	$N_1 = 262.7 \frac{kg}{batch}$	$N_2 = 262.7 \frac{kg}{batch}$

$Total_Input := N_1$

$Total_Output := N_2$

$Total_Input = 262.7 \frac{kg}{batch}$

$Total_Output = 262.7 \frac{kg}{batch}$

APPENDIX B

ENERGY BALANCE

Appendix B entails an illustration of energy balance calculations using MathCad numerical software with an example of the surge drum other equipments follow the same procedure.

Define: Stream indices in order 1-LVGO, 2-HVGO, 3-$W_{Distillate}$ 4-C_2, 5-C_3, 6-C_4, 7-Gasoline, 8-LCO, 9-DCO, 10-Slurry, 11-Catalyst

Stream indices $j := 1..16$ $k := 1..4$

Species indices $s := 1..11$

Index of the heat capacity coefficients

Table 1: Heat Capacity Coefficients

$$\begin{pmatrix} \text{LVGO} \\ \text{HVGO} \\ W_{\text{Distillate}} \\ C_2 \\ C_3 \\ C_4 \\ \text{Gasoline} \\ \text{LCO} \\ \text{DCO} \\ \text{Slurry} \\ \text{Catalyst} \end{pmatrix} \quad c := \begin{pmatrix} 23124 & 0 & 0 & 0 & 0 \\ 12341 & 0 & 0 & 0 & 0 \\ 321121 & 0 & 0 & 0 & 0 \\ -1.7675 & 1.1429 & -3.2361\cdot10^{-4} & 4.2431\cdot10^{-4} & 0 \\ 3.9489\cdot10^{1} & 3.951\cdot10^{1} & 2.1141\cdot10^{-3} & 3.9649\cdot10^{-7} & 0 \\ 3.0903\cdot10^{1} & 1.533\cdot10^{-1} & 2.6348\cdot10^{-3} & 7.2723\cdot10^{-8} & 0 \\ 6.425\cdot10^{1} & -1.318\cdot10^{-1} & 3.541\cdot10^{-3} & -1.3332\cdot10^{-6} & 0 \\ 7.4513\cdot10^{1} & -9.6697\cdot10^{-2} & 3.4765\cdot10^{-3} & -1.3212\cdot10^{-6} & 0 \\ 4.9787\cdot10^{-9} & -6.52\cdot10^{-2} & 3.4029\cdot10^{-3} & -1.2535\cdot10^{-6} & 0 \\ 4.1085\cdot10^{-9} & -5.3706\cdot10^{-2} & 3.3714\cdot10^{-3} & -1.2366\cdot10^{-6} & 0 \\ 43145.6 & 0 & 0 & 0 & 0 \end{pmatrix}$$

Reference Temperature $T_r := 298\cdot K$

Stream Number $n := 1$

Stream Temperature $T_n := 303\cdot K$

Enthalpy Equation for reacting components:

$$h(s,T_r) = \left[h_{f_s} + \left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array} \right] \cdot \frac{kJ}{kmol} \right] \cdot m_{s,j}$$

$$h(s,T_r) = \left[h_{f_s} + \left[\begin{array}{l} c_{s,1}\cdot\left(\frac{T}{K} - \frac{T_{ref}}{K}\right) + c_{s,2}\cdot\left(\frac{T^2}{K^2} - \frac{T_{ref}{}^2}{K^2}\right) + c_{s,3}\cdot\left(\frac{T^3}{K^3} - \frac{T_{ref}{}^3}{K^3}\right) \ldots \\ + c_{s,4}\cdot\left(\frac{T^4}{K^4} - \frac{T_{ref}{}^4}{K^4}\right) \end{array} \right] \cdot \frac{kJ}{kmol} \right] \cdot m_{s,j}$$

Enthalpy Equation for non-reacting components:

$$h(s,T_r) = \left[\left[\begin{array}{l} c_{s,1}\cdot(T - T_{ref}) + c_{s,2}\cdot\left(T^2 - T_{ref}{}^2\right) + c_{s,3}\cdot\left(T^3 - T_{ref}{}^3\right) \ldots \\ + c_{s,4}\cdot\left(T^4 - T_{ref}{}^4\right) \end{array} \right] \cdot \frac{kJ}{kmol} \right] \cdot m_{s,j}$$

$$h(s,T_r) = \left[\left[\begin{array}{l} c_{s,1}\cdot\left(\frac{T}{K} - \frac{T_{ref}}{K}\right) + c_{s,2}\cdot\left(\frac{T^2}{K^2} - \frac{T_{ref}{}^2}{K^2}\right) + c_{s,3}\cdot\left(\frac{T^3}{K^3} - \frac{T_{ref}{}^3}{K^3}\right) \ldots \\ + c_{s,4}\cdot\left(\frac{T^4}{K^4} - \frac{T_{ref}{}^4}{K^4}\right) \end{array} \right] \cdot \frac{kJ}{kmol} \right] \cdot m_{s,j}$$

Unit 1: Energy Balance of FCC Feed Surge Drum D-01

Enthalpy of stream 1:

$T_r := 25°C$ $T_1 := 26°C$ $T := T_1$

$$H_{s,1} := z_{s,1} \cdot \left[\left[\left[\frac{c_{s,1}}{1} \cdot \left(\frac{T}{K} - \frac{T_r}{K} \right) + \frac{c_{s,2}}{2} \cdot \left[\left(\frac{T}{K} \right)^2 - \left(\frac{T_r}{K} \right)^2 \right] \ldots \right] \right] \cdot \frac{J}{kmol} \right]$$
$$+ \frac{c_{s,3}}{3} \cdot \left[\left(\frac{T}{K} \right)^3 - \left(\frac{T_r}{K} \right)^3 \right] + \frac{c_{s,4}}{4} \cdot \left[\left(\frac{T}{K} \right)^4 - \left(\frac{T_r}{K} \right)^4 \right]$$

$H_{s,1} =$

LVGO	12.105	$\cdot \frac{kJ}{batch}$
HVGO	4.687	
W Distillate	60.389	
C_2	0	
C_3	0	
C_4	0	
Gasoline	0	
LCO	0	
DCO	0	
Slurry	0	
Catalyst		

Enthalpy of stream 1 is:

$$\sum_s H_{s,1} = 77.181 \frac{kJ}{batch}$$

Enthalpy of stream 2:

$T_r := 25°C$ $T_2 := 25.5°C$ $T := T_2$

$$H_{s,2} := z_{s,2} \cdot \left[\left[\left[\frac{c_{s,1}}{1} \cdot \left(\frac{T}{K} - \frac{T_r}{K} \right) + \frac{c_{s,2}}{2} \cdot \left[\left(\frac{T}{K} \right)^2 - \left(\frac{T_r}{K} \right)^2 \right] \dots \right. \right. \right. \left. \left. \left. + \frac{c_{s,3}}{3} \cdot \left[\left(\frac{T}{K} \right)^3 - \left(\frac{T_r}{K} \right)^3 \right] + \frac{c_{s,4}}{4} \cdot \left[\left(\frac{T}{K} \right)^4 - \left(\frac{T_r}{K} \right)^4 \right] \right] \right] \cdot \frac{J}{kmol} \right]$$

$H_{s,2} =$

LVGO	6.053	$\cdot \frac{kJ}{batch}$
HVGO	2.344	
W	30.194	
Distillate	0	
C_2	0	
C_3	0	
C_4	0	
Gasoline	0	
LCO	0	
DCO	0	
Slurry	0	
Catalyst		

Enthalpy of stream 2 is:

$$\sum_s H_{s,2} = 38.591 \frac{kJ}{batch}$$

Inlet Enthalpy:

$$\Delta H_{in_1} := \sum_s H_{s,1}$$

$$\Delta H_{in_1} = 77.181 \frac{kJ}{batch}$$

Outlet Enthalpy:

$$\Delta H_{out_1} := \sum_s H_{s,2} \qquad \Delta H_{out_1} = 38.591 \frac{kJ}{batch}$$

Heat load:

$$Duty_1 := \Delta H_{out_1} - \Delta H_{in_1}$$

$$Duty_1 = -38.591 \frac{kJ}{batch}$$

APPENDIX C

Appendix C entails an illustration of equipment design calculations using MathCad numerical tool with an example of the surge drum other equipments follow the same procedure.

Equipment Design of FCC Unit Feed Tank D-01

Flow rate of the feed to the Storage tank

$F_a := N_1$ $\quad F_a = 262.7 \frac{kg}{batch}$

Density of the feed to the tank

$\rho_a := 1056 \frac{kg}{m^3}$

Volumetric flow rate of the feed to tank

$V_3 := \frac{F_a}{\rho_a}$

$V_3 = 0.249 \frac{m^3}{batch}$

Storage time, τ:

$\tau := 5 \cdot hr$

$\tau = 18000 s$

Volume of the tank required for the storage time:

$V := V_3 \cdot \tau$ $\quad V = 0.138 m^3$

Allowing for 30% safety allowance:

$$V := \frac{V}{0.7}$$

$$Q_{Storage_Tank1} := V$$

$$V = 0.197 m^3$$

The rule of thumb states that the maximum volume of the tank is 20 m^3, the number of tank required is:

$$NO := \frac{V}{1 \cdot m^3}$$

The optimum proportions for a cylindrical container. A classical example of the optimization of a simple function.

The surface are, A, of a closed cylinder is:

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

Where D = vessel diameter

L = vessel length (or height)

This will be the objective function which is to be minimized; simplified:

$$f(D \times L) = D \times L + \frac{D^2}{2}$$

For a given volume, V, the diameter and length are related by:

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

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

and the objective function becomes

$$f(D) = 4\frac{v}{\pi D} + \frac{D^2}{2}$$

Setting the differential of this function to zero will give the optimum value for D

$$-4\frac{V}{\pi D^2} + D = 0$$

$$D := \sqrt[3]{4\frac{V}{\pi}}$$

$$D = \sqrt[3]{4\frac{119.632}{3.142}}$$

$D = 0.631\text{m}$ $DD_1 := D$

From literature, Dinert et al., 1993 ratio of tank diameter to tank height (D/H) is:

$$Ratio = \frac{D}{H} = 0.4$$

$$Ratio := 0.6$$

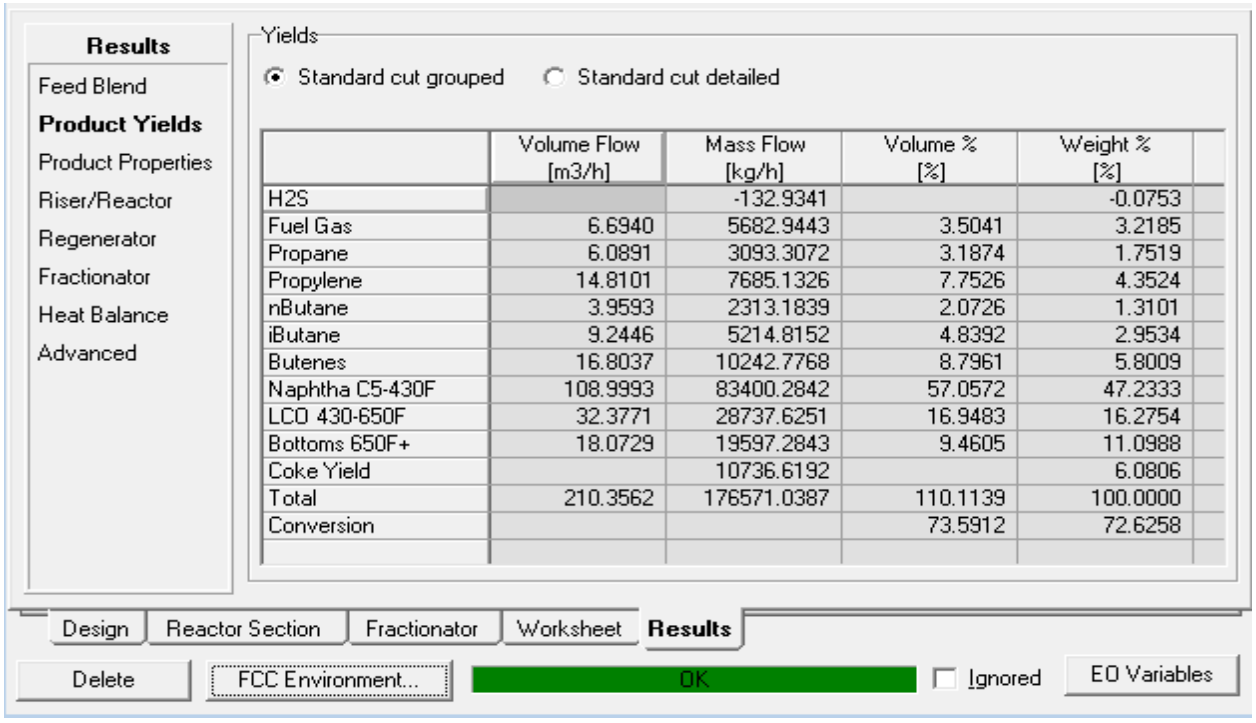
$$H := \frac{D}{Ratio}$$

$H = 1.002\,m$ $\qquad$ $HH_1 := H$

APPENDIX D

Computer Simulation Details showing screen shots of Hysys simulation results.

Table D1: Product Distribution



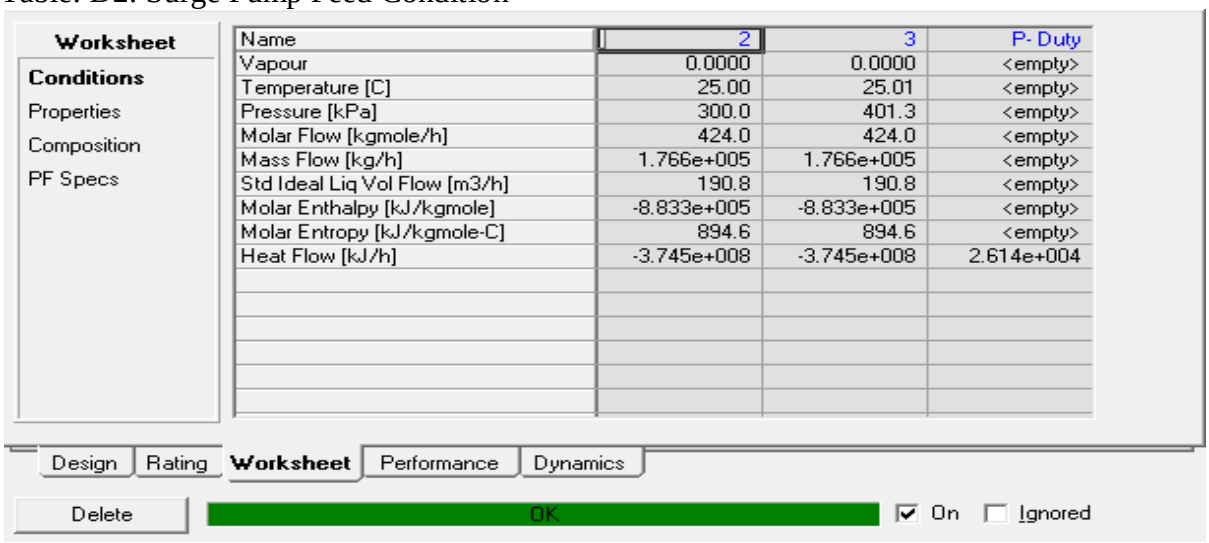
	Volume Flow [m3/h]	Mass Flow [kg/h]	Volume % [%]	Weight % [%]
H2S		-132.9341		-0.0753
Fuel Gas	6.6940	5682.9443	3.5041	3.2185
Propane	6.0891	3093.3072	3.1874	1.7519
Propylene	14.8101	7685.1326	7.7526	4.3524
nButane	3.9593	2313.1839	2.0726	1.3101
iButane	9.2446	5214.8152	4.8392	2.9534
Butenes	16.8037	10242.7768	8.7961	5.8009
Naphtha C5-430F	108.9993	83400.2842	57.0572	47.2333
LCO 430-650F	32.3771	28737.6251	16.9483	16.2754
Bottoms 650F+	18.0729	19597.2843	9.4605	11.0988
Coke Yield		10736.6192		6.0806
Total	210.3562	176571.0387	110.1139	100.0000
Conversion			73.5912	72.6258

Major Equipment Specifications based on Hysys Computer Simulation

Equipment specification base on the simulation result is presented in tables and figures below:

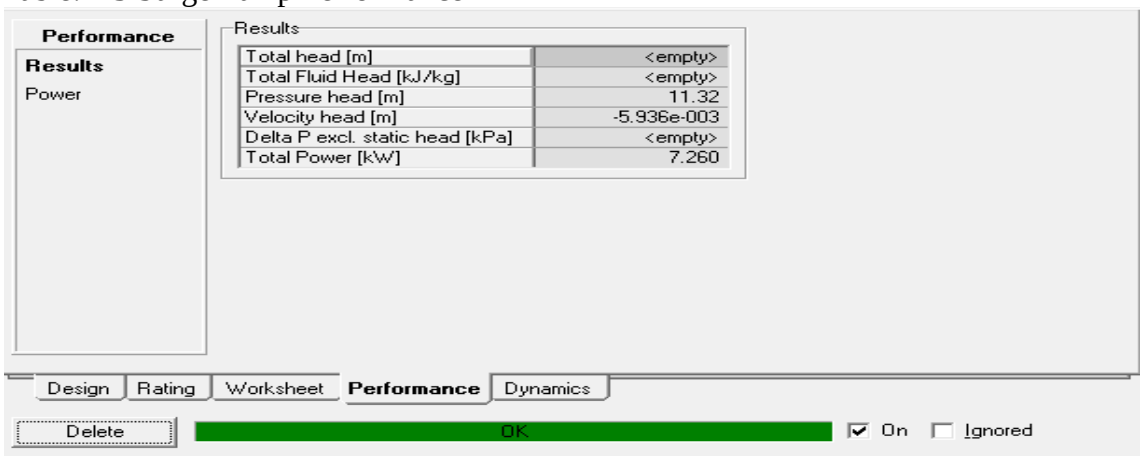
Surge Pump (P-01)

Table: D2: Surge Pump Feed Condition



Name	2	3	P-Duty
Vapour	0.0000	0.0000	<empty>
Temperature [C]	25.00	25.01	<empty>
Pressure [kPa]	300.0	401.3	<empty>
Molar Flow [kgmole/h]	424.0	424.0	<empty>
Mass Flow [kg/h]	1.766e+005	1.766e+005	<empty>
Std Ideal Liq Vol Flow [m3/h]	190.8	190.8	<empty>
Molar Enthalpy [kJ/kgmole]	-8.833e+005	-8.833e+005	<empty>
Molar Entropy [kJ/kgmole-C]	894.6	894.6	<empty>
Heat Flow [kJ/h]	-3.745e+008	-3.745e+008	2.614e+004

Table: D3 Surge Pump Performance



Heat Exchanger (E-01)

Table D4: Exchanger Feed Condition

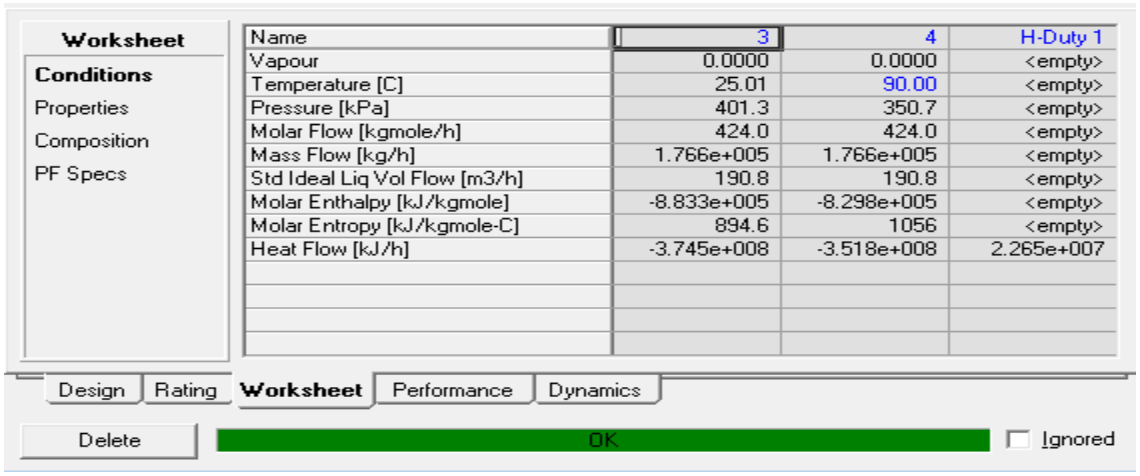
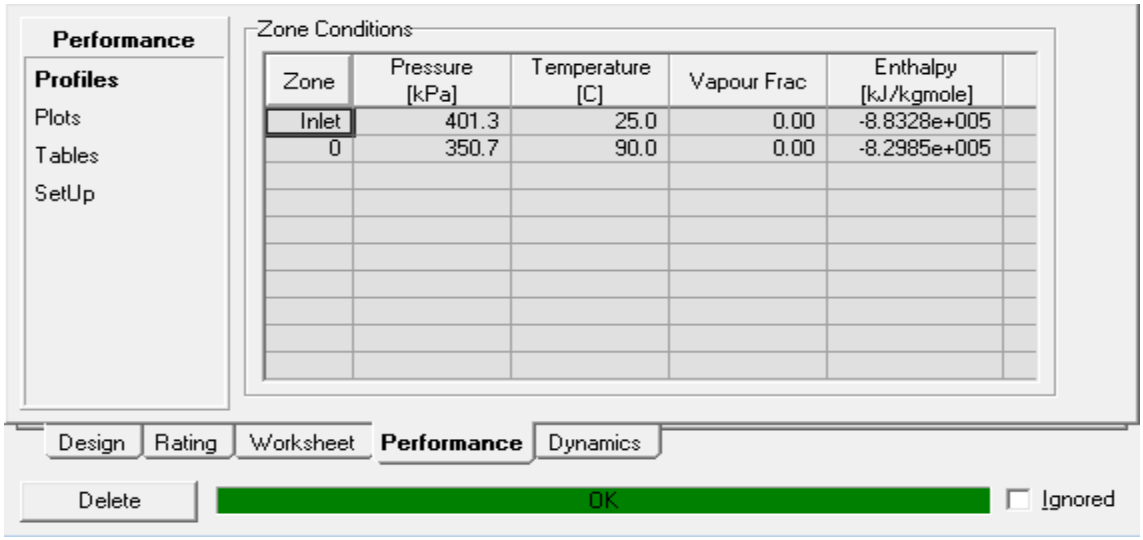
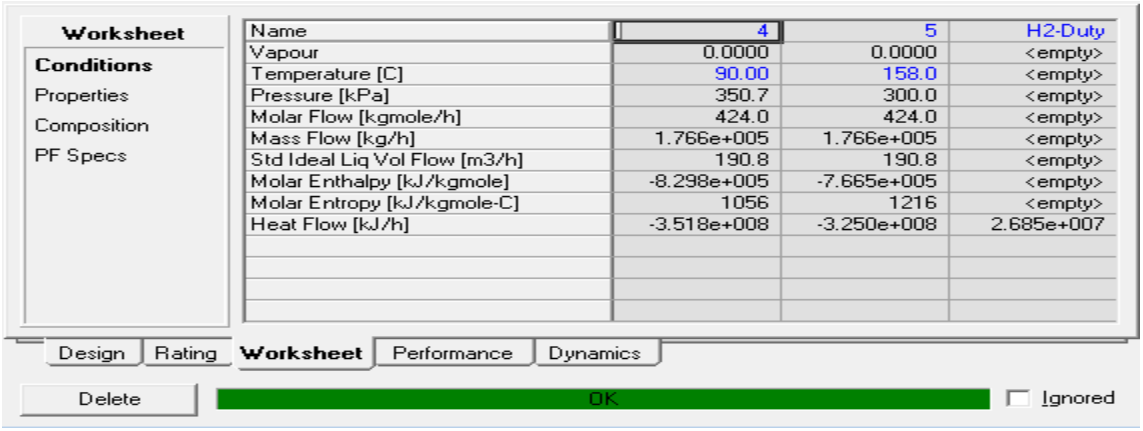


Table D5: Exchanger Performance



Heat Exchanger (E-02)

Table D6: Exchanger Feed Condition



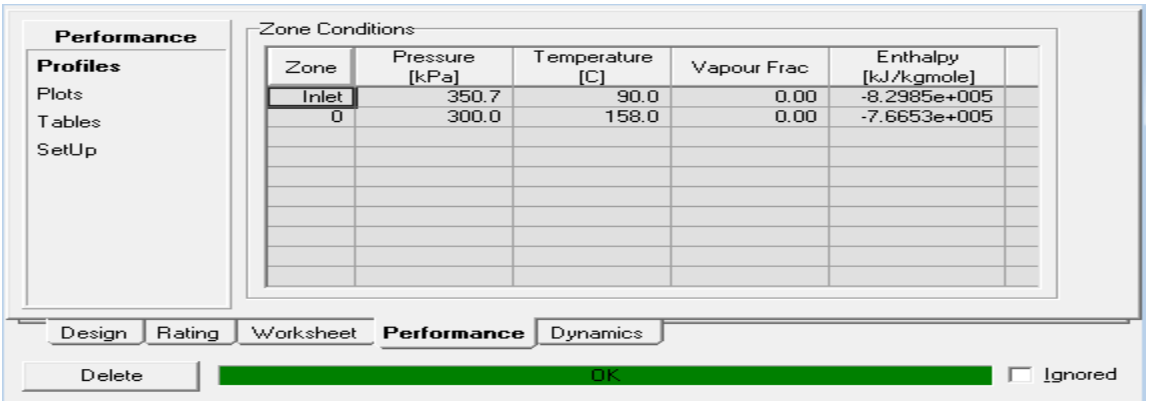
Worksheet
Conditions
Properties
Composition
PF Specs

Name	4	5	H2-Duty
Vapour	0.0000	0.0000	<empty>
Temperature [C]	90.00	158.0	<empty>
Pressure [kPa]	350.7	300.0	<empty>
Molar Flow [kgmole/h]	424.0	424.0	<empty>
Mass Flow [kg/h]	1.766e+005	1.766e+005	<empty>
Std Ideal Liq Vol Flow [m3/h]	190.8	190.8	<empty>
Molar Enthalpy [kJ/kgmole]	-8.298e+005	-7.665e+005	<empty>
Molar Entropy [kJ/kgmole-C]	1056	1216	<empty>
Heat Flow [kJ/h]	-3.518e+008	-3.250e+008	2.685e+007

Design | Rating | Worksheet | Performance | Dynamics

Delete | OK | Ignored

Table D7: Exchanger Performance



Performance
Profiles
Plots
Tables
SetUp

Zone Conditions

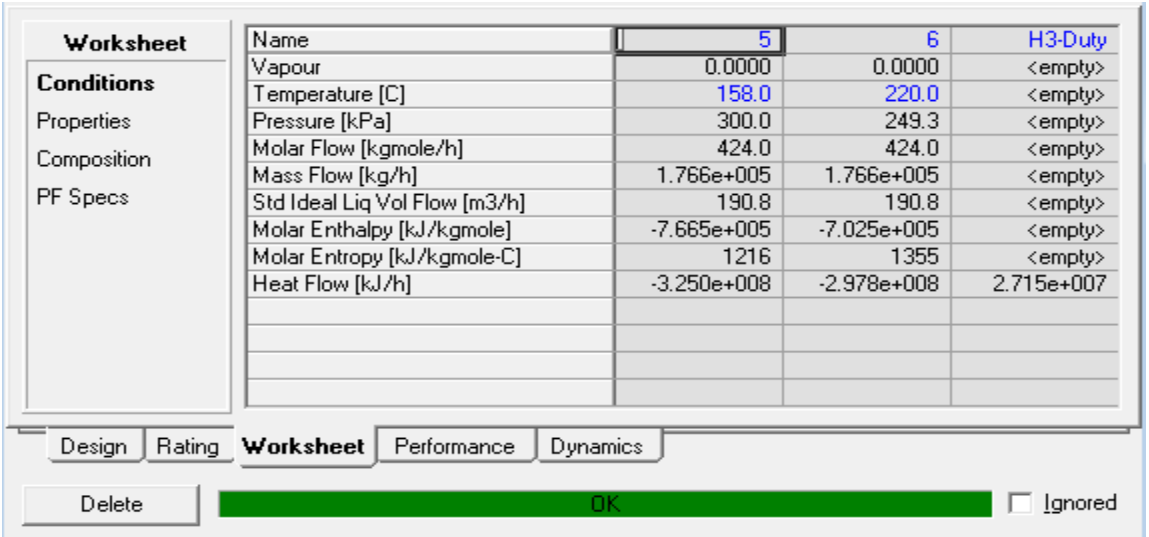
Zone	Pressure [kPa]	Temperature [C]	Vapour Frac	Enthalpy [kJ/kgmole]
Inlet	350.7	90.0	0.00	-8.2985e+005
0	300.0	158.0	0.00	-7.6653e+005

Design | Rating | Worksheet | Performance | Dynamics

Delete | OK | Ignored

Heat Exchanger (E-03)

Table D8: Exchanger Feed Condition



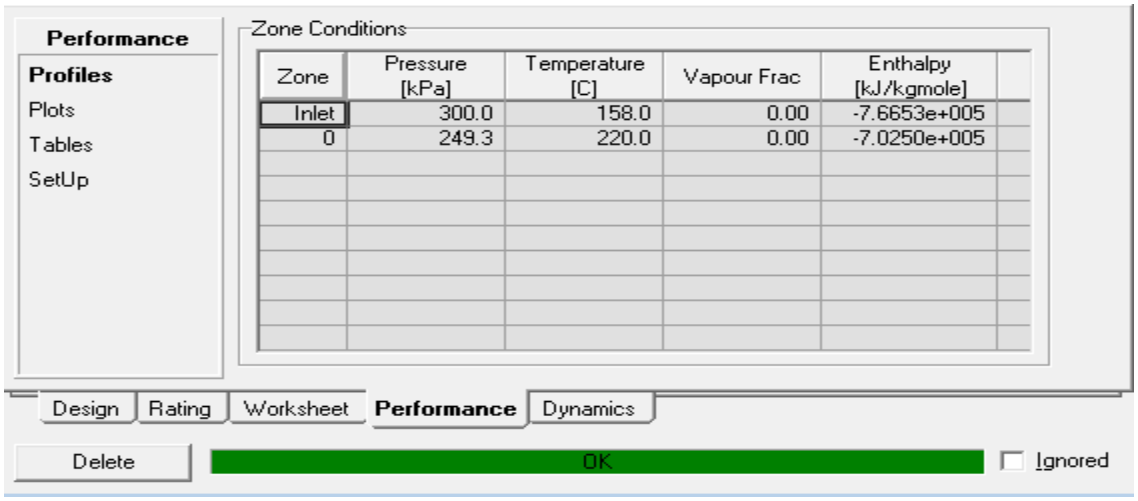
Worksheet
Conditions
Properties
Composition
PF Specs

Name	5	6	H3-Duty
Vapour	0.0000	0.0000	<empty>
Temperature [C]	158.0	220.0	<empty>
Pressure [kPa]	300.0	249.3	<empty>
Molar Flow [kgmole/h]	424.0	424.0	<empty>
Mass Flow [kg/h]	1.766e+005	1.766e+005	<empty>
Std Ideal Liq Vol Flow [m3/h]	190.8	190.8	<empty>
Molar Enthalpy [kJ/kgmole]	-7.665e+005	-7.025e+005	<empty>
Molar Entropy [kJ/kgmole-C]	1216	1355	<empty>
Heat Flow [kJ/h]	-3.250e+008	-2.978e+008	2.715e+007

Design | Rating | Worksheet | Performance | Dynamics

Delete | OK | Ignored

Table D9: Exchanger Performance



Feed Heater

Table D10: Heater Feed Condition

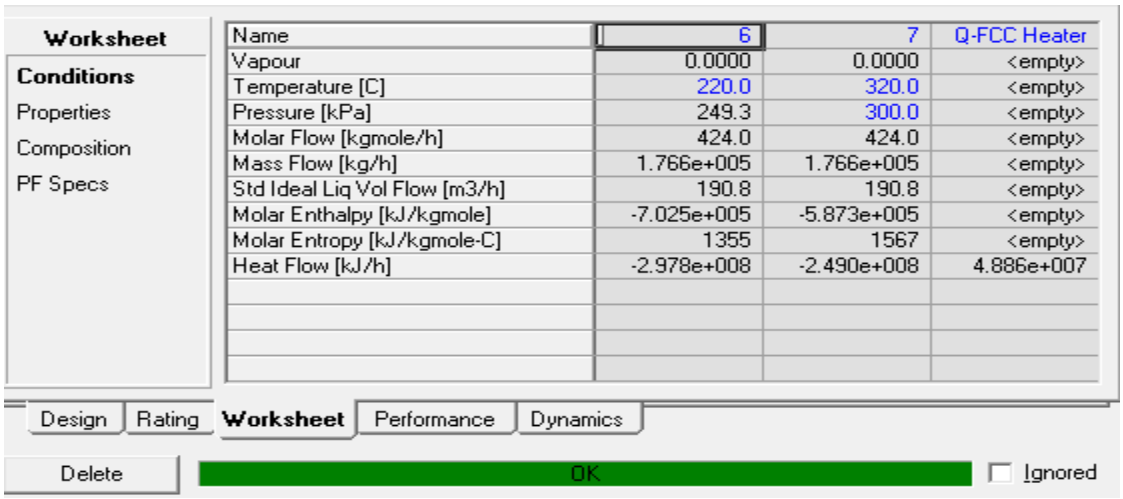


Table D11: Heater Specs

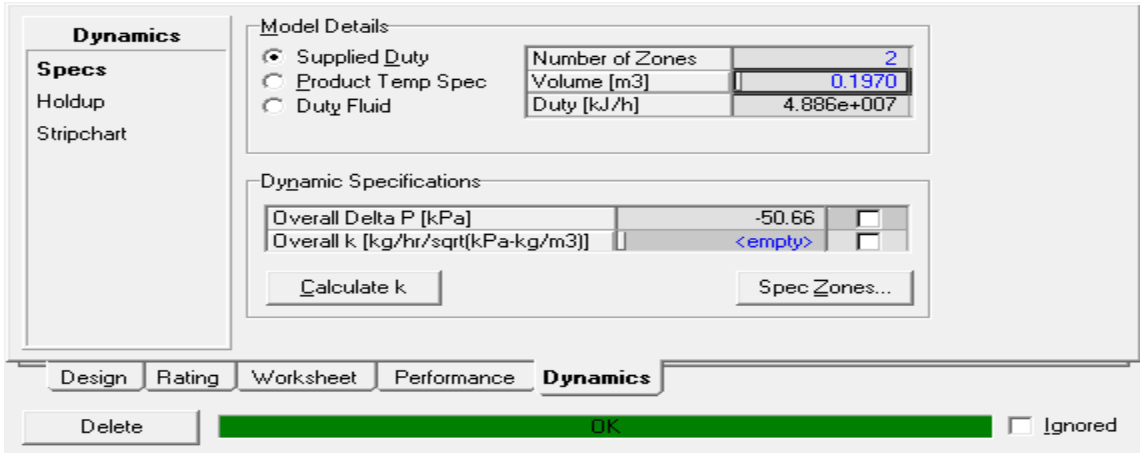
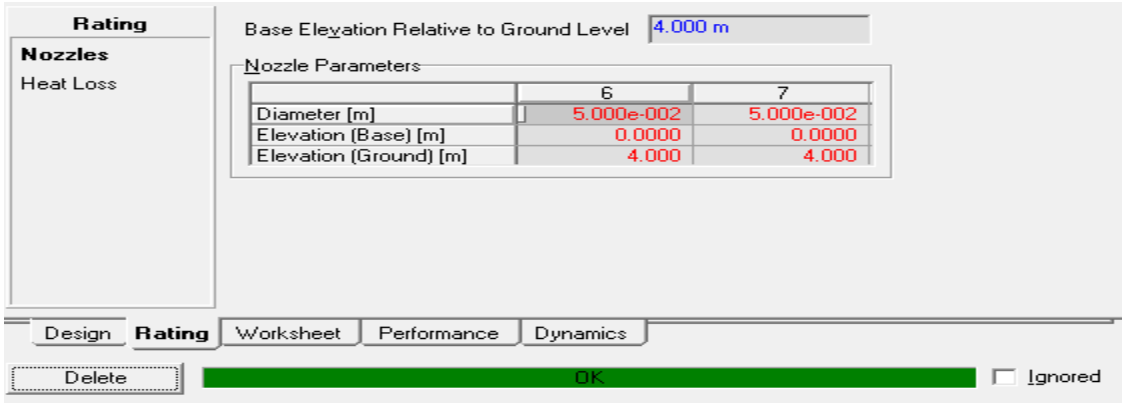


Table D12: Heater Rating



FCC Reactor

Table D13: Reactor Feed Condition

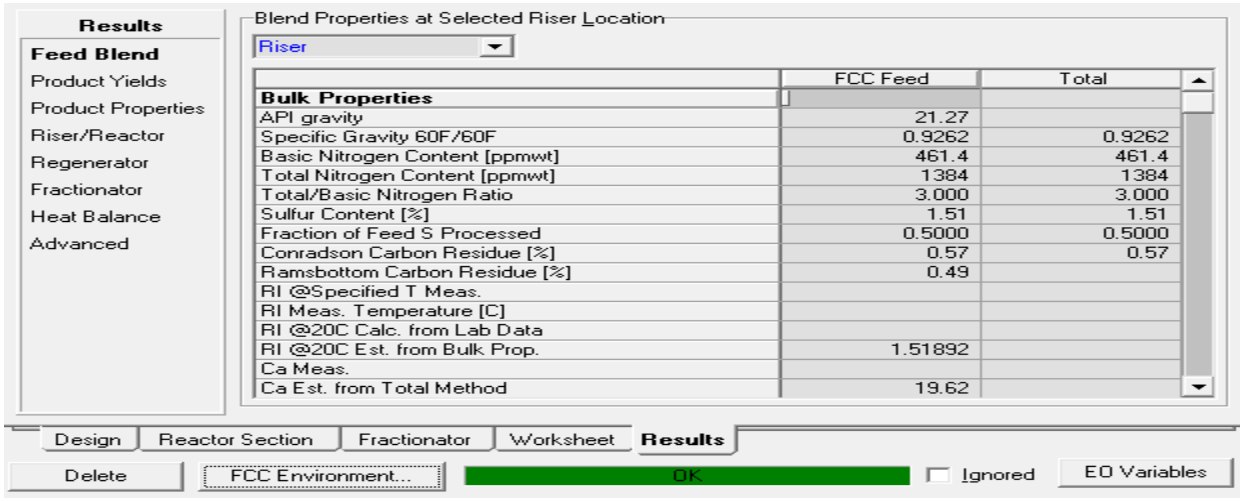


Table D14: Reactor Operating Condition

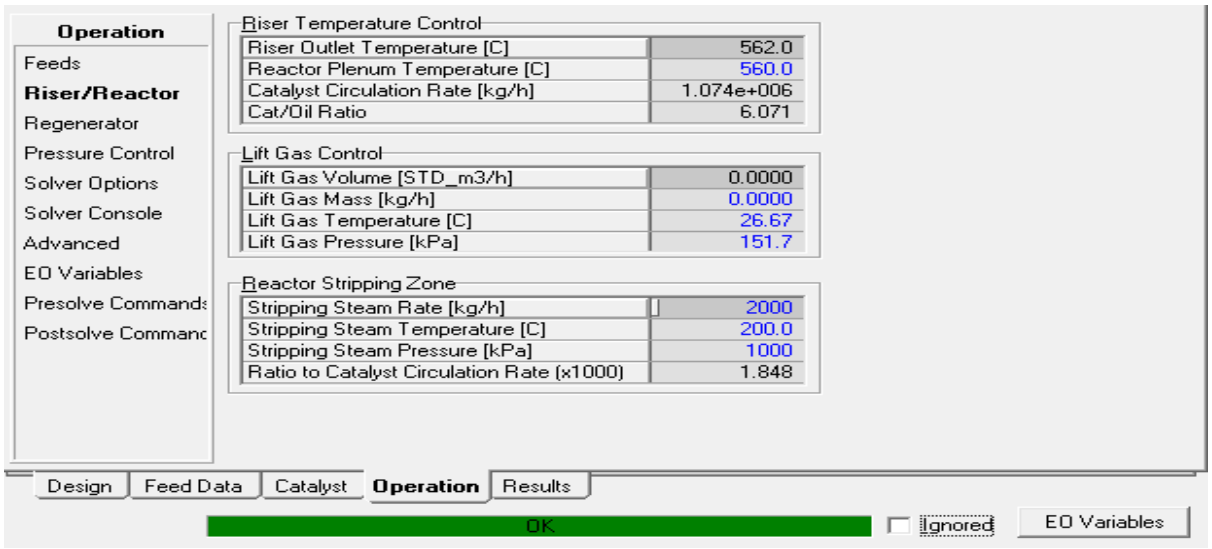
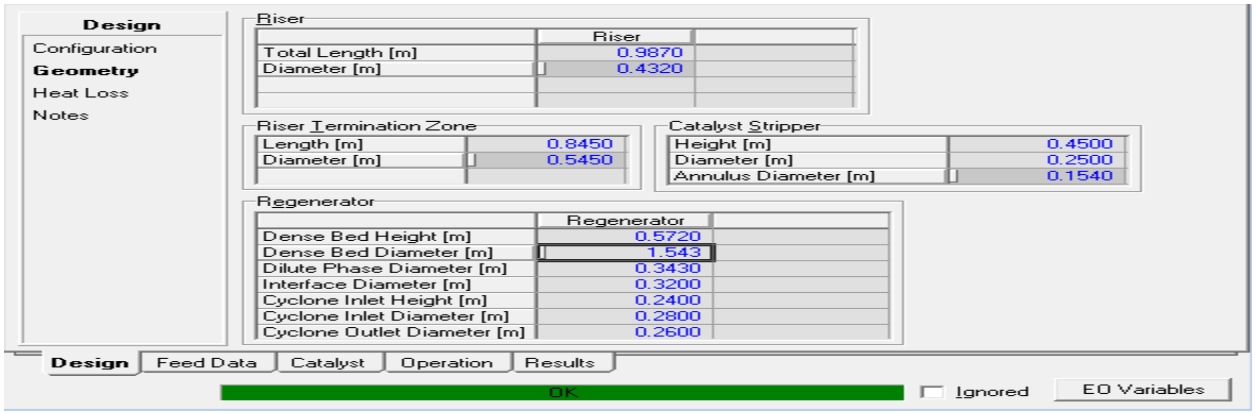


Table D15: Reactor Geometry



Distillation Column

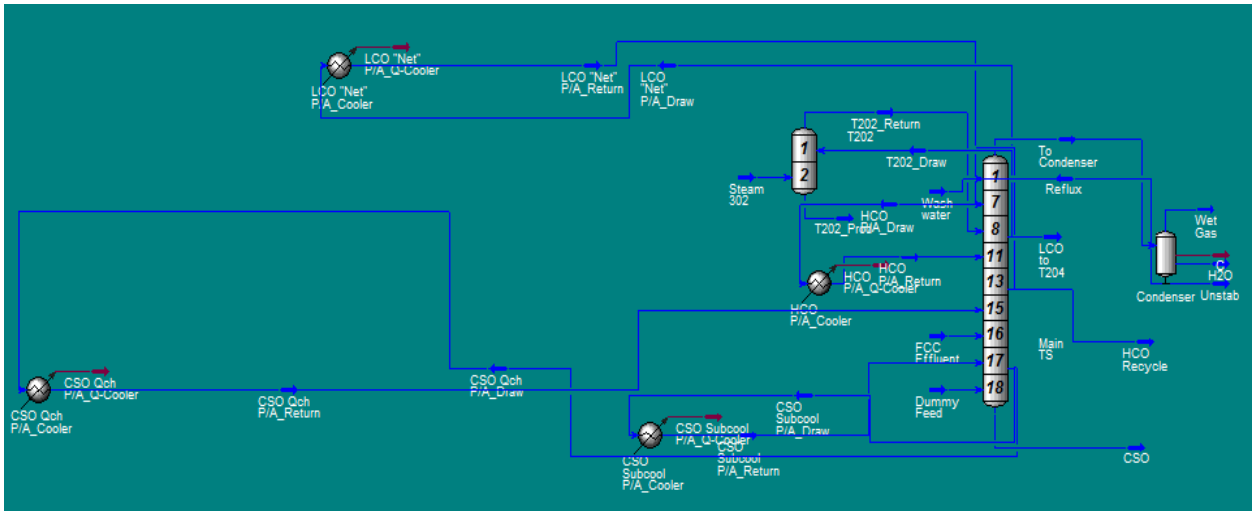


Figure D1: Column Stages

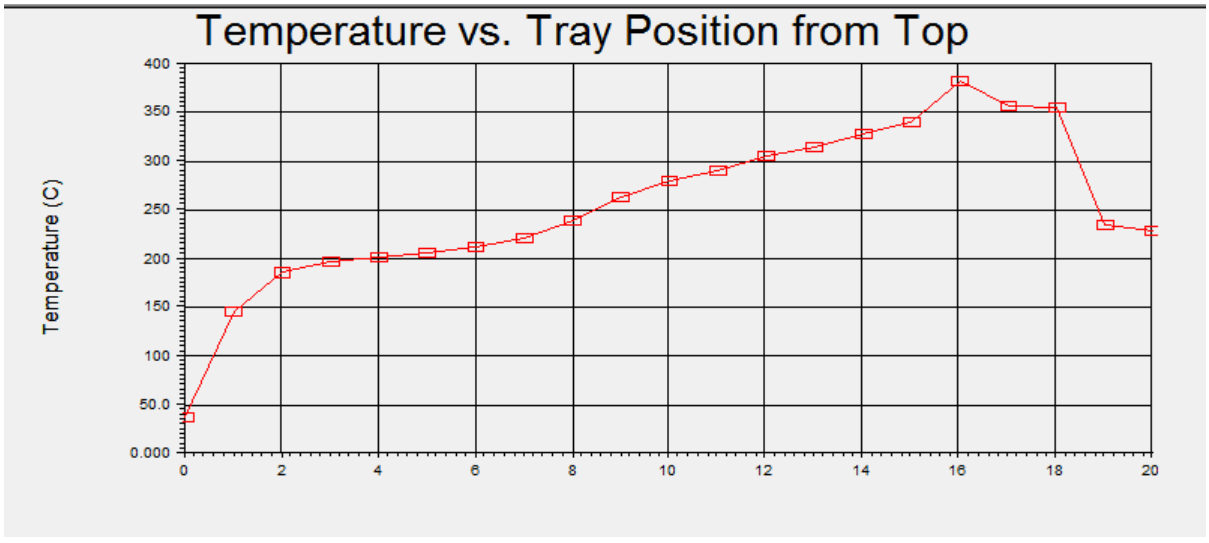
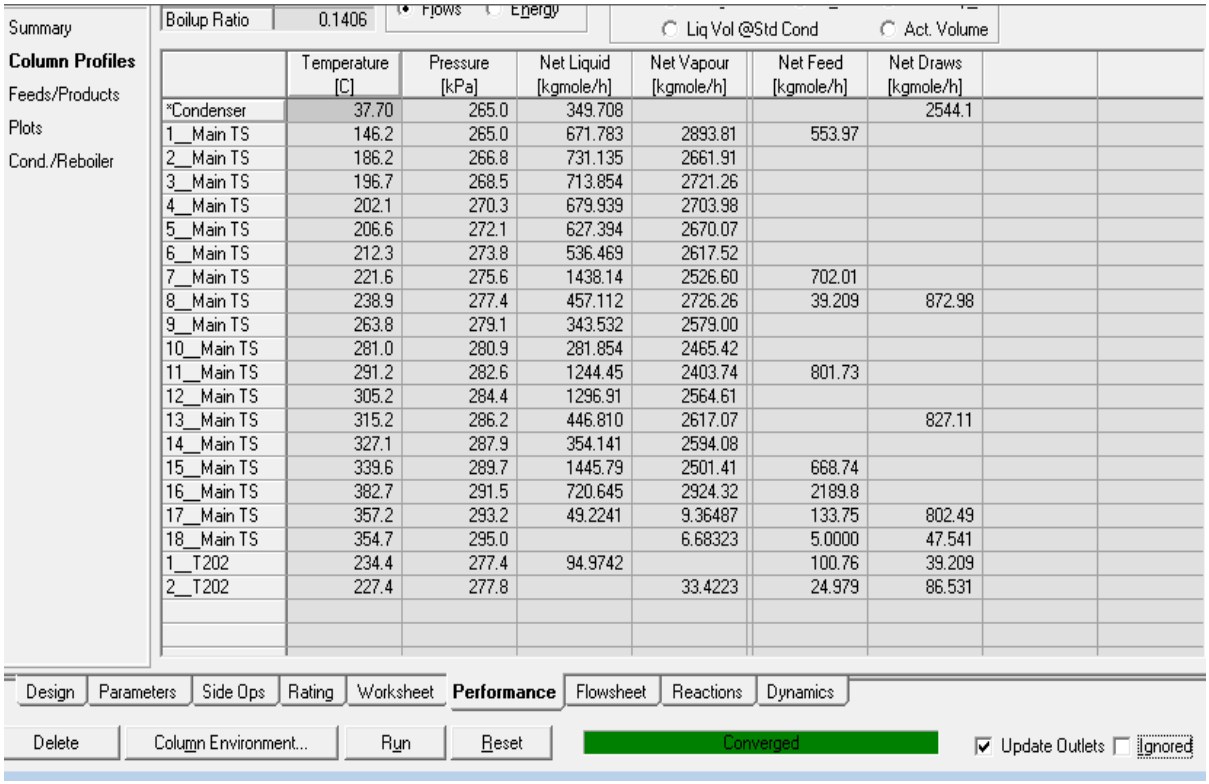


Figure D2 Temperature Profile

Table D16: Column Profiles



Summary
Column Profiles
Feeds/Products
Plots
Cond./Reboiler

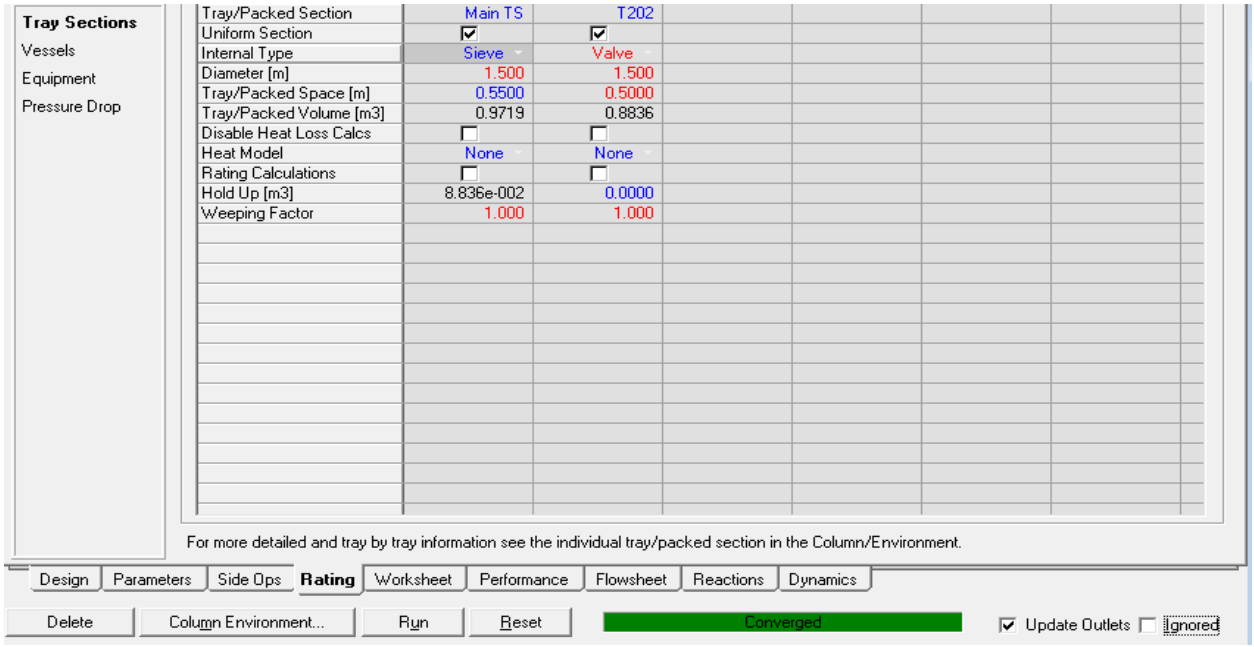
Boilup Ratio 0.1406 Flows Energy Liq Vol @Std Cond Act. Volume

	Temperature [C]	Pressure [kPa]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]	Net Feed [kgmole/h]	Net Draws [kgmole/h]
*Condenser	37.70	265.0	349.708			2544.1
1__Main TS	146.2	265.0	671.783	2893.81	553.97	
2__Main TS	186.2	266.8	731.135	2661.91		
3__Main TS	196.7	268.5	713.854	2721.26		
4__Main TS	202.1	270.3	679.939	2703.98		
5__Main TS	206.6	272.1	627.394	2670.07		
6__Main TS	212.3	273.8	536.469	2617.52		
7__Main TS	221.6	275.6	1438.14	2526.60	702.01	
8__Main TS	238.9	277.4	457.112	2726.26	39.209	872.98
9__Main TS	263.8	279.1	343.532	2579.00		
10__Main TS	281.0	280.9	281.854	2465.42		
11__Main TS	291.2	282.6	1244.45	2403.74	801.73	
12__Main TS	305.2	284.4	1296.91	2564.61		
13__Main TS	315.2	286.2	446.810	2617.07		827.11
14__Main TS	327.1	287.9	354.141	2594.08		
15__Main TS	339.6	289.7	1445.79	2501.41	668.74	
16__Main TS	382.7	291.5	720.645	2924.32	2189.8	
17__Main TS	357.2	293.2	49.2241	9.36487	133.75	802.49
18__Main TS	354.7	295.0		6.68323	5.0000	47.541
1__T202	234.4	277.4	94.9742		100.76	39.209
2__T202	227.4	277.8		33.4223	24.979	86.531

Design | Parameters | Side Ops | Rating | Worksheet | **Performance** | Flowsheet | Reactions | Dynamics

Delete | Column Environment... | Run | Reset | Converged | Update Outlets | Ignored

Table D17: Column Tray Sections



Tray Sections
Vessels
Equipment
Pressure Drop

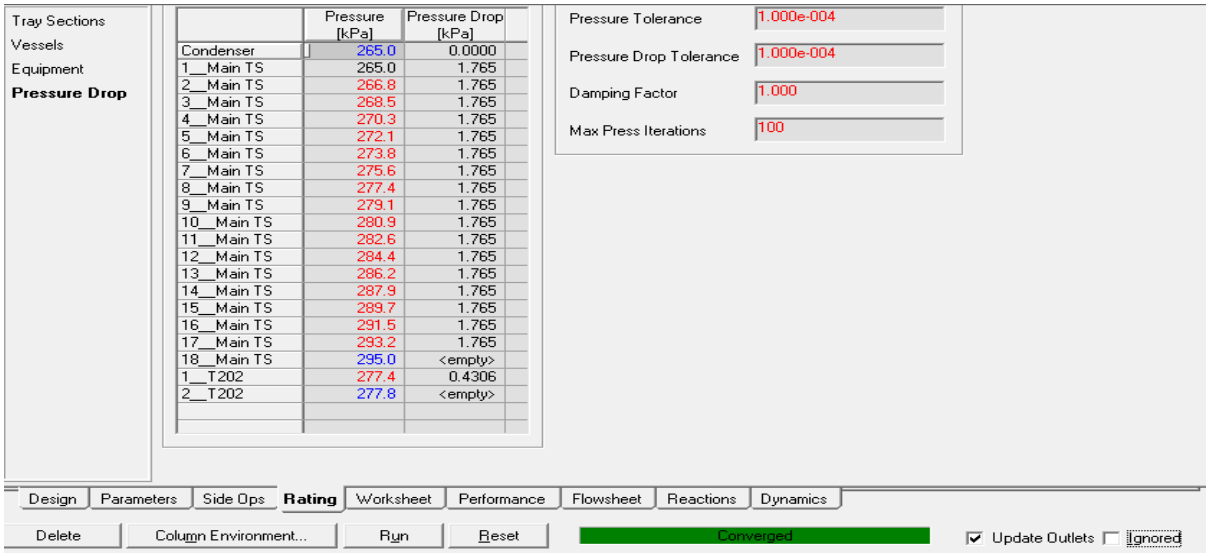
Tray/Packed Section	Main TS	T202
Uniform Section	☑	☑
Internal Type	Sieve	Valve
Diameter [m]	1.500	1.500
Tray/Packed Space [m]	0.5500	0.5000
Tray/Packed Volume [m3]	0.9719	0.8836
Disable Heat Loss Calcs	☐	☐
Heat Model	None	None
Rating Calculations	☐	☐
Hold Up [m3]	8.836e-002	0.0000
Weeping Factor	1.000	1.000

For more detailed and tray by tray information see the individual tray/packed section in the Column/Environment.

Design | Parameters | Side Ops | **Rating** | Worksheet | Performance | Flowsheet | Reactions | Dynamics

Delete | Column Environment... | Run | Reset | Converged | Update Outlets | Ignored

Table D18: Column Pressure Drop



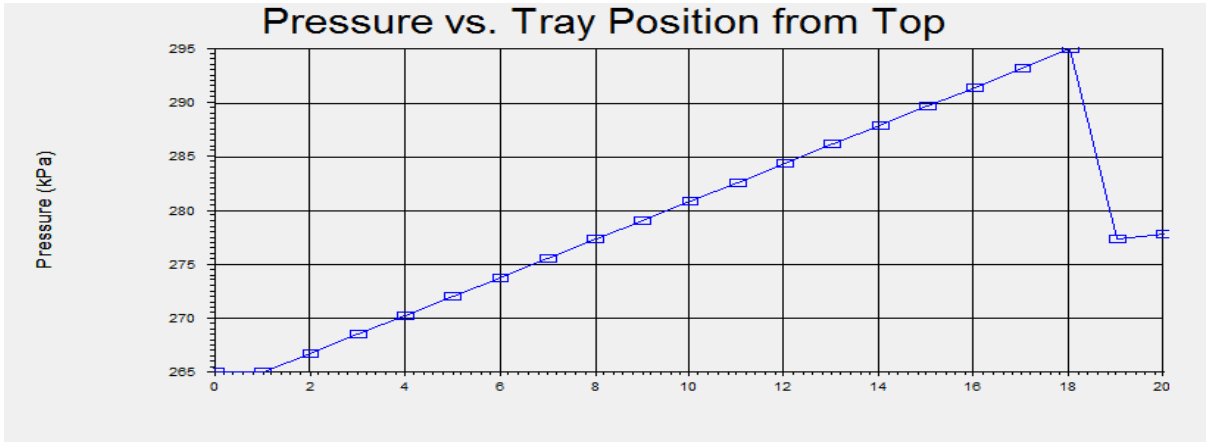


Figure D3 Pressure Profile

Overhead Condenser

Table D19: Condenser Specs

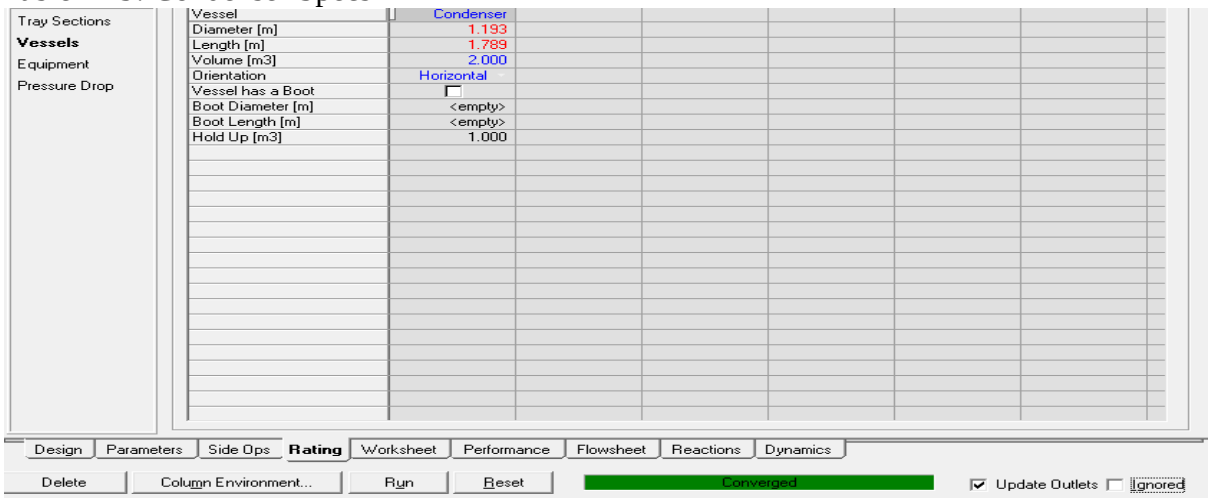
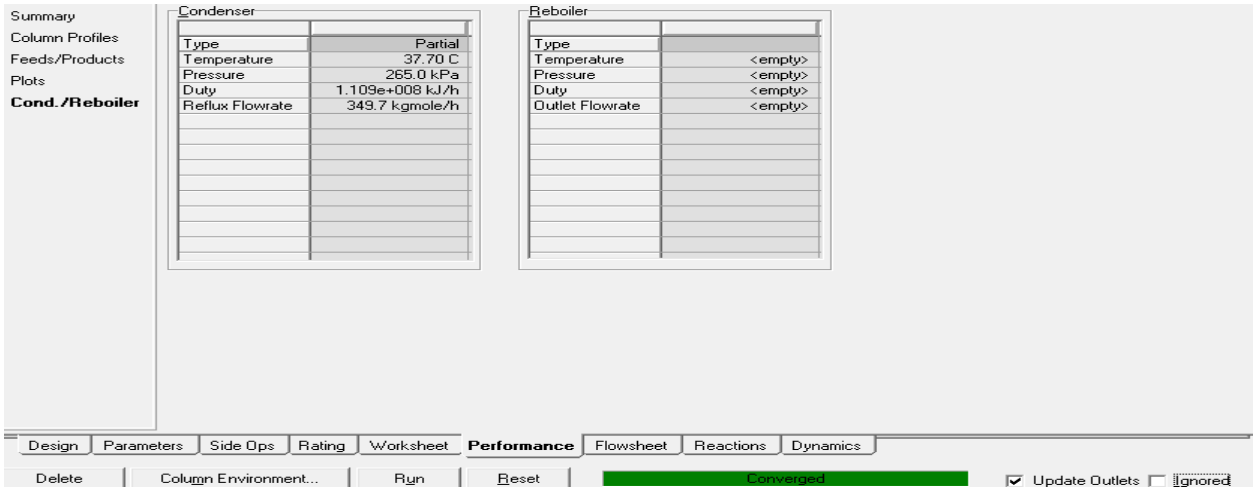


Table D20: Condenser Condition



Overhead Separator

Table D21: Overhead Separator Condition

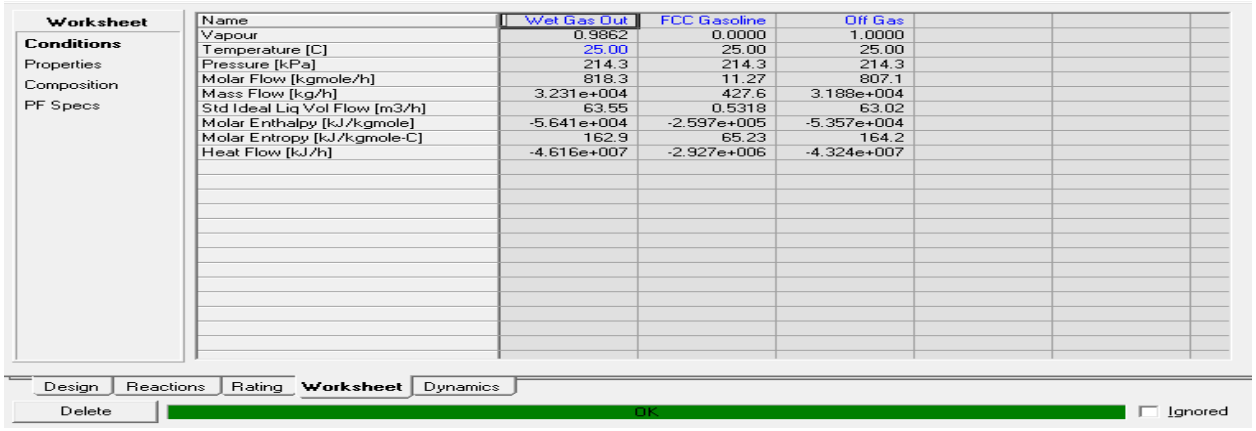


Table D22: Overhead Separator Sizing

