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تأثير الهواء الساخن علي إعاده تنشيط العامل الحفاز في وحده التكسير الحفزي

EFFECT OF HOT AIR ON REACTIVATION OF CATALYST IN THE FLUIDIZED CATALYTIC CARACKING PROCESS (RFCC)

A thesis submitted in partial fulfillment of the requirements for the degree of f B.SC. Honors in petroleum and Refining engineering

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ألأية

قال تعالى:

{شهد الله انه لا إله إلا هو والملائكة وألو العلم قائما بالقسط لا إله إلا هو العزيز الحكيم }

[آلعمران: الآية 18]

Dedication

We would like to dedicate this research to our Dear parents.

To the people who paved our way of science and knowledge.

To the taste of the most beautiful moments with our friends.

Also we dedicate this research to our brothers and sisters who spent their life to provide sufficient, Happiness and love.

Finally we dedicate this research to our colleagues In Transportation and refining department,

To all batch 23th petroleum engineering Students. And all batch 5th refining engineering.

Acknowledgment

أرشدنا النبي صلي عليه وسلم الي مكافأه من أحسن النيات فإن لم تجد فإننا ندعو جزاك الله خيرا.

روى ابو داوود(1672)عن عبدالله بن عمر رضي الله عنه قال:قال رسول الله صلي الله عليه وسلم (من صنع اليكم معروفا فكافئوه فإن لم تجدوا ما تكافئونه فادعو له حتي تروا أنكم قد كافأتموه).

صححه الالباني في إرواءالغليد(1617).

First of all, we would like to thank Great and Almighty God who gave us the power to fulfill this study. Thanks should be given to our supervisor’s Dr. Mohammed Mustafa A. A. Nasir and Mr.Adil Abbas and Eng. RAMAH AZHRY and ENG.HOSSAM ELDIN MOSTAFA.

Thank you all

Abstract

The objective of this study is assess effect of air amount in regenerative process in Fluid catalytic cracking unit (FCCU by using ASPEN HYSYS simulation software.

The results showed that increase the amount of air present in the regenerator caused an increase in the regeneration of catalyst.

This study has shown the optimum amount to achieve carbon percentage of less than 0.05% on regeneration catalyst is **3400** Nm^3/min .

التجـــريد

الهدف من هذه الدراسه تقييم تأثير كميه الهواء في عمليه التجديد في وحده التكسير الحفزي بإستخدام برنامج

(ASPEN HYSYS) .

وأظهرت الدراسه أن زياده كميه الهواء في وحده التجديد تؤدي الي زياده نشاطيه العامل الحفاز.

وقد أظهرت هذه الدراسه أن كميه المثلي لتحقيق نسبه كربون أقل من 0.05% عند التجديد هي $3400 Nm^3$ /min.

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SYMBOLS

SYMBOLS	Description	Unit
$Mcat$	Catalytic circulation	Kg/hr.
ρ_c	Catalyst density	Kg/ m^3
ρ_b	Bulk density of feed	Kg/m^3
V	volumetric flow rate of feed	m^3/hr.
Vav	the average volumetric flow rate of feed in the riser,	m^3/s
Vo	the volumetric flow rate at the out let of the riser	m^3/s
Vi	the volumetric flow rate at the inlet of the riser	m^3/s
Uav	the average linear velocity	m/s
Uo	the velocity at the inlet of the riser	m/s
Ui	the average at the outlet of the riser	m/s
A	cross section area	m^2
D	riser diameter	m
H	riser height	m
Uav	Average linear velocity	m/s
τ	Residence time in the riser	s
ΔP	Pressure drop	N/m^2
U_g^2	gas velocity through the riser	m/s
g_c	gravitantial conversion factor,	kg.m/s^2.N
W_S	solid flux	$kg/m^2.s$
U_P	partial velocity	m/s
V	volume of the settler	m^3
Do	diameter of riser	m
Di	diameter of settler	m
L	height of settler	m
ρ	catalyst density in settler	kg/m^3
Vca	volume of catalyst + coke	m^3
D_d	dense phase diameter	m
L_F	the fluidized bed height	m
m_{cat}	the mass flow rate of the catalyst	kg

t	the optimum residence time	min
W_{Ca}	the catalyst weight	kg
P	the density of the catalyst particles	Kg/m³
P1	- Dilute phase pressure	MPa
T1	Dilute phase temperature	k
T	thickness	m
Di	internal diameter	m
J	welded joint factor;	
Pi	internal pressure MPa	MPa
F	stress of material	N/mm²
M_{fg}	Mass flue gas	kg
M_{rca}	Mass of Regan. catalyst	kg
M_{SCa}	Mass of spent catalyst	kg
M_{air}	Mass of air	kg
M_{p}	Mass of product	kg
M_{so}	Mass of steam outlet	kg
M_{F}	Mass of feed	kg
M_{SI}	Mass of steam inlet	kg
FCC	fluid catalytic cracking	

Chapter (1)

1.1 Introduction

Global demand for transportation fuels will continue to grow and this demand will be met largely by gasoline and diesel fuels. The fluid catalytic cracking (FCC) process continues to play a key role in an integrated refinery as the primary conversion process of crude oil to lighter products. In the next two decades, the FCC process will be likely used for biofuels and possibly for reducing CO2 emissions. For many refiners, the cat cracker is the key to profitability because the successful operation of the unit determines whether or not the refiner can remain 3
966competitive in today's market. (Sadeghbeigi, 2012)

The fluid catalytic cracking (FCC) unit is the essential transformation unit done numerous refineries and it is one of the most important and complex processes in the petroleum refining industry. It converts heavy material feeds consisting of high boiling points like gas oil into lighter and more valuable products like gasoline, liquefied petroleum gas (LPG), and olefins by using a zeolite catalyst and it helps to produce about half of the total gasoline output in a refinery. 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. The heavy vacuum gas oil (VGO) has a boiling point range of 343- 565°C and an average molecular weight ranging from about 200 to 600 or higher. A wide range of feed stocks can be processed in the FCC unit such as cracked gas oils, deasphalted oils and hydro treated gas oil. (Ateya, 2016)

The modern FCC units are all continuous processes which operate 24 hours a day for as much as two to three years between shutdowns for routine maintenance. (Gary, 2001) The primary purpose of the **FCC** process has been to product gasoline, distillate, and C3/C4 olefin from low-value excess refinery gas oils and heavier refinery stream in this unit we use ZYULITE action to break the bonds. (Sadeghbeigi, 2012)

The modern RFCC unit can accept a broad range of feedstocks, a fact which contributes to RFCC's reputation as one of the most flexible refining processes in use today.

Examples of common feedstocks for conventional distillate feed FCC units are:

- Atmospheric gas oils
- Vacuum gas oils
- Thermally cracked gas oils
- Solvent deasphalted oils
- Lube extracts
- Hydrocracker bottoms

Products from the FCC and RFCC processes are typically as follows:

- Fuel gas (ethane and lighter hydrocarbons)
- C3 and C4 liquefied petroleum gas (LPG)
- Gasoline
- Light cycle oil (LCO)
- Fractionator bottoms (slurry oil)
- Coke (combusted in regenerator)
- Hydrogen Sulfide (from amine regeneration)

1.2 Problem Statement:

When cracking reactions occur, reduction in catalyst activity occurs due to carbon deposition on the catalytic surface and the decrease production rate.

1.3 Objectives:

1- Effect of hot air on reactivtion catalyst in the fluidized catalytic caracking process

2- Increasing amount of gasoline production rate.

1.4. Scope:

1- Material balance around (Reactor and regenerator).
2- Simulation to unit by ASPEN HYSYS .
3- Design unit (reactor and regeneration).

Chapter (2)
Literature Review

2.1 Historical perspective about (FCC) UNIT:

Catalytic cracking was first commercialized in 1936 by Eugene Hurdy.

The fixed bed process was a major improvement over the thermal cracking processes it replaced due to the improved yield distribution and superior product properties. Multiple vessels were utilized that alternated between cracking, stripping, regeneration, and purge cycles. This configuration was quickly replaced by a moving bed reactor and a separate regenerator or kiln that first used a bucket lift to move the pelleted catalyst followed later by a pneumatic air lift system. The last of these units was built around 1960. (JONES†, 2006)

In 1936, a new cracking process opened the way to higher-octane gasoline; this process was catalytic cracking. The process is basically the same as thermal cracking but differs by the use of a catalyst, which is not (in theory) consumed in the process and directs the course of the cracking reactions to produce more of the desired higher-octane hydrocarbon products. (Speight, 2007)

Standard Oil of New Jersey developed their own cracking process rather than pay the large royalty being asked at the time.

They commercialized the fluid catalytic cracking (FCC) process in three years, starting in 1939 and culminating in 1942 with the start-up of PCLA#1 at their Baton Rouge, Louisiana refinery. The inherent superiority of the fluid process to transfer both heat and catalyst ultimately made it the catalytic cracking process of choice. (JONES†, 2006)

There has been continuous up gradation in catalytic cracking process from its incept of fixed bed technology to letter fluid bed catalytic cracking (FCC):

The fluid catalytic cracking (FCC) process is the heart of a modern refinery oriented toward maximum gasoline production. Within the entire refinery process, this process offers the greatest potential for increasing profitability; even a small improvement giving higher gasoline yields can result in a substantial economic gain. The FCC process increases the H/C ratio by carbon rejection in a continuous process and is used to convert the high - boiling, high - molecular - weight hydrocarbon fractions (typically, a blend of heavy straight -Run gas oil, light vacuum gas oil, and heavy vacuum gas oil) to more valuable gasoline, olefin gases, and other products. (John Wiley & Sons, 2011)

2.2 Flow diagram of (FCC) Unit:

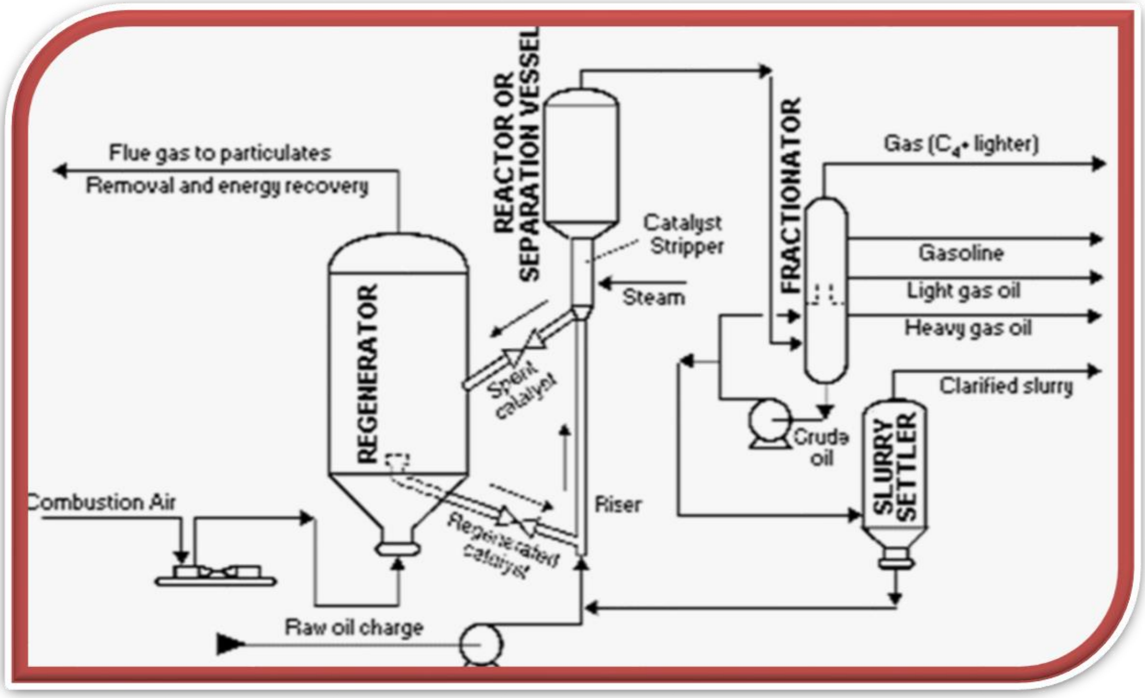


Fig. (2. 1) Schematic of the fluid catalytic cracking unit

2.3Reactor:

Preheated fresh feedstock, plus any recycle feed, is charged to the base of the riser reactor. Upon contact with hot regenerated catalyst, the feedstock is vaporized and converted to lower-boiling fractions (light cycle oil, gasoline, C3 and C4 LPG, and dry gas).

Product vapors are separated from spent catalyst in the disengage cyclones and flow via the overhead line to the main fractionators and vapor recovery unit for quenching and fractionation. Coke formed during the cracking reactions is deposited on the catalyst, thereby reducing its activity. The coked catalyst, which is separated from the reactor products in the disengage cyclones, flows via the stripper and spent catalyst standpipe to the regenerator.

The discharge rate from the standpipe is controlled by the spent catalyst plug valve. In the regenerator, coke is removed from the spent catalyst by combustion with air. Air is supplied to the regenerator air distributor s from an air blower. Flue gas from the combustion of coke exits the regenerator through two-stage cyclones which remove all but a trace of catalyst from the flue gas. Flue gas is collected in an external plenum chamber and flows to the flue gas train. Regenerated catalyst, with its activity restored, is returned to the riser via the regenerated catalyst plug valve (MEYERS, 2002).

2.3.1 Major Components of the Reactor:

2.3.1.1 ATOMAX Feed Injection System

The catalyst then enters a feed injection cone surrounded by multiple, flat-spray, atomizing feed injection nozzles. The flat, fan-shaped sprays provide uniform coverage and maximum penetration of feedstock into catalyst, and prevent catalyst from bypassing feed in the injection zone. Proprietary feed injection nozzles, known as ATOMAX nozzles, are used to achieve the desired feed atomization and spray pattern while minimizing feed pressure requirements. The hot regenerated catalyst vaporizes the oil feed, raises it to reaction temperature, and supplies the necessary heat for cracking. The cracking reaction proceeds as the catalyst and vapour mixture flow up the riser. The riser outlet temperature is controlled by the amount of catalyst admitted to the riser by the catalyst plug valve. (MEYERS, 2002)

2.3.1.2 Riser Quench:

The riser quench system consists of a series of nozzles uniformly spaced around the upper section of riser. A portion of the feed or a recycle stream from the main fractionators is injected through the nozzles into the riser to rapidly reduce the temperature of the riser contents. The heat required to vaporize the quench is supplied by increased fresh feed preheat or by increased catalyst circulation. This effectively increases the temperature in the lower section of the riser above that which would be achieved in anon-quenched operation, thereby increasing the vaporization of heavy feeds, increasing gasoline yield, olefin production and gasoline octane.

2.3.1.3 Spent Catalyst Stripping:-

Catalyst separated in the cyclones flows through the respective dip legs and discharges into the stripper bed. In the stripper, hydrocarbon vapors from within and around the catalyst particles are displaced by steam into the disengage dilute phase, minimizing hydrocarbon carry-under with the spent catalyst to the regenerator. Stripping is a very important function because it minimizes regenerator bed temperature and regenerator air requirements, resulting in increased conversion in regenerator temperature or air-limited operations. (MEYERS, 2002)

The catalyst entering the stripper is contacted by up flowing steam introduced through two steam distributors. The majority of the hydrocarbon vapors entrained with the catalyst are displaced in the upper stripper bed. The catalyst then flows down through a set of hat and doughnut baffles. In the baffled section, a combination of residence time and steam partial pressure is used to allow the hydrocarbons to diffuse out of the catalyst pores into the steam introduced via the lower distributor. (MEYERS, 2002) Stripped catalyst, with essentially all strippable hydrocarbons removed, passes into a standpipe, which is aerated with steam to maintain smooth flow. At the base of the standpipe, a plug valve regulates the flow of catalyst to maintain the spent catalyst level in the stripper. (MEYERS, 2002).

2.4 Regenerator-Heat/Catalyst Recovery:

The spent catalyst entering the regenerator usually contains between 0.5 and 1.5 wt. % coke. Components of coke are carbon, hydrogen, and trace amounts of sulphur and organic nitrogen molecules. These components burn according to the reactions given in Table (2.1).

2.4.1The regenerator has three main functions:

1. It restores catalyst activity.

2. It supplies heat for cracking reactions.

3. It delivers fluidized catalyst to the feed nozzles.

Air provides oxygen for the combustion of this 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 fluidized state.

The air/oxygen enters the regenerator through an air distribution system (Figure 2.2) located near the bottom of the regenerator vessel. The design of the air distributor is important in achieving efficient and reliable catalyst regeneration. (Speight, 2007)

Table (2. 1) The Main reaction of the coke Combustion reaction

	kcal/kg of C, H2, or S	BTU/lb. of C, H2, or S
$C + 0.5O_2 \rightarrow CO$	2200	3968
$CO + 0.5O_2 \rightarrow CO_2$	5600	10100
$C + O_2 \rightarrow CO_2$	7800	14100

$H_2 + 0.5O_2 \rightarrow H_2O$	**28900**	**52125**
$S + xO \rightarrow SO_X$	**2209**	**3968**
$N + xO \rightarrow NO_X$	**-**	**-**

Air distributors are often designed for a 1.0- to 2.0-psi (7_15kPa) pressure drop to ensure positive air flow through all nozzles.

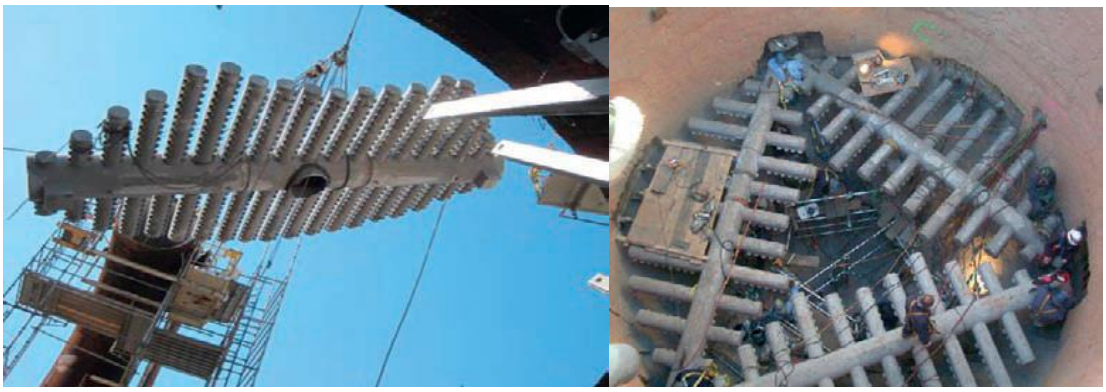


Fig. (2. 2) Examples of air distributor designs (Speight, 2007)

2.4.2Partial Versus Complete Combustion:

Two distinctly different modes of regeneration are practiced partial combustion and complete combustion.

Complete combustion generates more energy and the coke yield is decreased; partial combustion generates less energy and the coke yield is increased. In complete combustion, the excess reaction component is oxygen, so more carbon generates more combustion. In partial combustion, the excess reaction component is carbon, all the oxygen is consumed, and an increase in coke yield means a shift from CO2 to CO.

FCC regeneration can be further subdivided into low intermediate-, and high-temperature regeneration. In low Temperature regeneration (about 1,190_F or 640_C), complete Combustion is impossible. One of the characteristics of low-temperature regeneration is. That at 1,190_F, all three components (O2, CO, and CO2) 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. (Speight, 2007)

In the early 1970s, high-temperature regeneration was developed. 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 and no CO or no excess oxygen and a variable quantity of CO.
If there was excess oxygen, the operation was in full burn. If there was excess CO, the operation was in partial burn.

With a properly designed air/spent catalyst distribution system and potential use of CO combustion promoter, the regeneration temperature could be reduced and still maintain full burn mode of catalyst regeneration.

(Table 2.2) contains a matrix summarizing various aspects of catalyst regeneration. Regeneration is either partial or complete at low, intermediate, or high temperatures. 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 (Speight, 2007)

2.5 Carbon concentration in the regenerated catalyst

The molecular sieve catalyst is very sensitive to the carbon concentration of the regenerated catalyst. The carbon concentration of the regenerated catalyst means the remaining coke content on the catalyst after regeneration. If the carbon content of the regeneration catalyst is too high, the activity and the selectivity of the molecular sieve catalyst will be decreased; therefore, the conversion yield will be decreased greatly, and the gasoline yield will also decrease, the bromine value will increase and the induction period will decrease. Adopting high-temperature regeneration can decrease the carbon concentration of the catalyst, normally, 0.1% carbon content decreasing, 2-3 units of the activity will be increased.

Table (2.2) Regeneration characteristic:

Operating Region Regenerator Combustion	Partial Combustion Mode	Full Combustion Mode
Low temperature (nominally 1,190F/640C)	**Stable (small afterburning); O2, CO, and CO2 in the flue gas.**	**Not achievable**
Intermediate temperature (nominally 1,275F/690C)	**Stable (with combustion promoter).**	**Stable with combustion promote**
High temperature (nominally 1,350F/730C)	**tends to have high carbon on regenerated catalyst Stable Operation.**	**Stable operation**

2.6.RFCC Catalysts:-

A catalyst is a substance that affects the rate of a reaction but emerges from the process unchanged. A catalyst usually changes a reaction rate by promoting a different molecular Path (“mechanism”) for the reaction (Fogler & F. Thurnau Professor , 2011).

Paralleling the significant improvements in RFCC unit design was a corresponding improvement in RFCC catalysts. The first catalysts used were ground-up amorphous silica alumina. Whether synthetic or naturally occurring, these catalysts suffered from low activity and poor stability relative to the catalysts available today. Additionally, they had poor fluidization characteristics. Often, fines had to be collected from the flue gas and returned to the unit to assist in maintaining smooth catalyst circulation. In 1946, spray-dried (micro spheroidal) synthetic silica-alumina catalysts were introduced. This type of catalyst, containing 10 to 13 percent alumina, was in general use until a more active and stable catalyst high in alumina (25 wt. % alumina) became available in the late 1950s. (Fogler & F. Thurnau Professor , 2011).

Zeolite is the key ingredient of the FCC catalyst. It provides product selectivity and much of the catalytic activity. The catalyst’s performance depends largely on the nature and quality of the zeolite. Understanding the zeolite structure, types, cracking mechanism, and properties is essential in choosing the “right” catalyst to produce the desired yields. (Speight, 2007) .

The resulting catalysts exhibited significantly higher activity and stability compared with catalysts available at the time. These crystalline catalysts were, and are, ideally suited for the short-contact-time riser cracking concept. Besides being more active, these materials are more selective toward gasoline production which is the most important refinery product compared to the initial amorphous type. (Speight, 2007) .

Table (2. 2) Comparison of Amorphous and Zeolite Catalysts

	Amorphous	Zeolite
Coke, wt.%	4	4
Conversion, vol.%	55	65
C_5+gasoline, vol%	38	51
C_4's, vol%	17	16
C3gas, wt.%	7	6

Because a catalytic reaction occurs at the fluid–solid interface, a large interfacial area is almost always essential in attaining a significant reaction rate. In many catalysts, this area is provided by an inner porous structure (i.e., the solid contains many fine pores, and the surface of these pores supplies the area needed for the high rate of reaction). Sometimes pores are so small that they will admit small molecules but prevent large ones from entering. Materials with this type of pore are called molecular sieves. (Vennema, n.d.)

It has a well-defined lattice structure. Its basic building blocks are silica and alumina tetrahedral (pyramids). Each tetrahedron (Figure4.1) consists of a silicon or aluminium atom at the centre of the tetrahedron, with oxygen atoms at the four corners. The pore diameter of nearly all of today's FCC zeolite is 8.0 angstroms (A °). These small openings, with an internal surface area of roughly 600 m^2/g. Zeolites employed in the manufacture of the FCC catalyst are synthetic versions of naturally occurring zeolites called faujasites.

The properties of the zeolite play a significant role in the overall performance of the catalyst. Understanding these properties increases our ability to predict catalyst response to changes in unit operation.

The zeolite must retain its catalytic properties under the hostile conditions of the FCC operation. The reactor/ regenerator environment can cause significant changes in chemical and structural composition of the zeolite. (Speight, 2007)

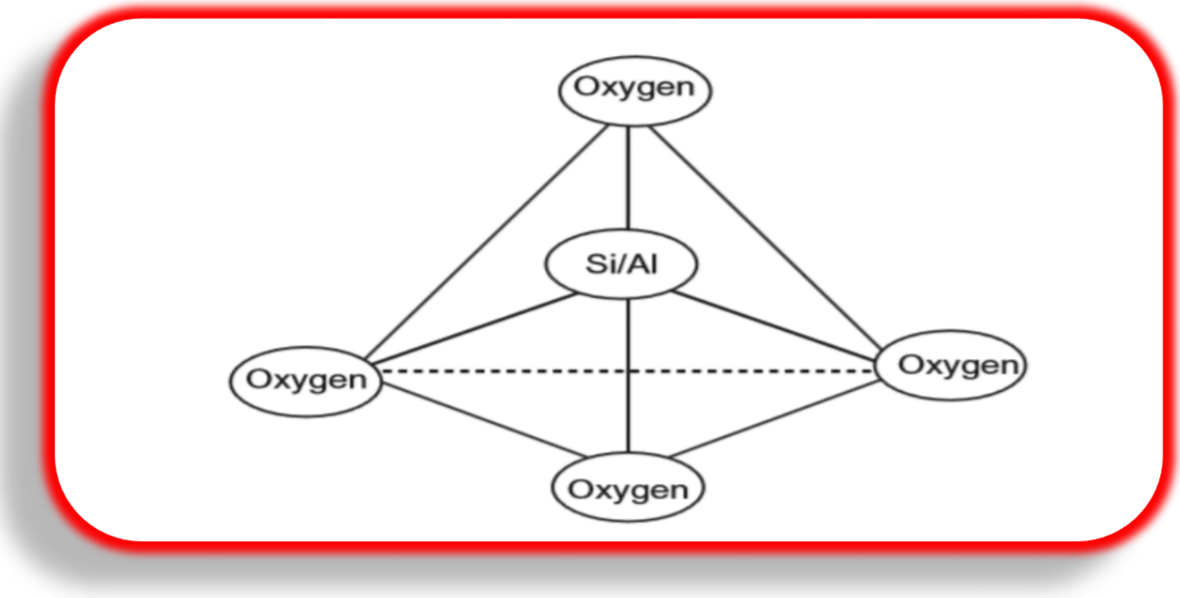


Fig. (2. 3) Silicon/aluminium-oxygen tetrahedron.

2.6.1 Additives:-

In some cases a catalyst consists of minute particles of an active material dispersed over a less active substance called a support. The active material is frequently a pure metal or metal alloy.

Such catalysts are called supported catalysts, as Distinguished from unsupported catalysts.

Analysts can also have small amounts of active ingredients added called promoters, which increase their activity. Examples of supported catalysts are the packed-bed catalytic converter in an automobile, the platinum-on-alumina catalyst used in petroleum reforming, and the vanadium pent oxide on silica used to oxidize sulfur dioxide in manufacturing sulphuric acid. (Speight, 2007).

2.6.2Catalyst Deactivation:-

Most catalysts used in the chemical and petrochemical industries are strongly affected by one or another form of deactivation, leading to poor performances and reduced life.

Deactivation of catalysts can involve sintering, poisoning or fouling, and the last two of these involve the adsorption/deposition of various species on the catalyst.

Catalyst poisons tend to be selective, since they deactivate by strong adsorption. The major purpose of the material herein is to link the more fundamental aspects of catalyst deactivation by poisoning.

To those related problems which plague industrial practitioners. Thus we are primarily concerned with events which occur on an intermediate size scale. Catalyst foul ants are much less selective, coke formation in zeolite catalysts is a spatially demanding reaction and, as such, is particularly sensitive to zeolite pore structure.

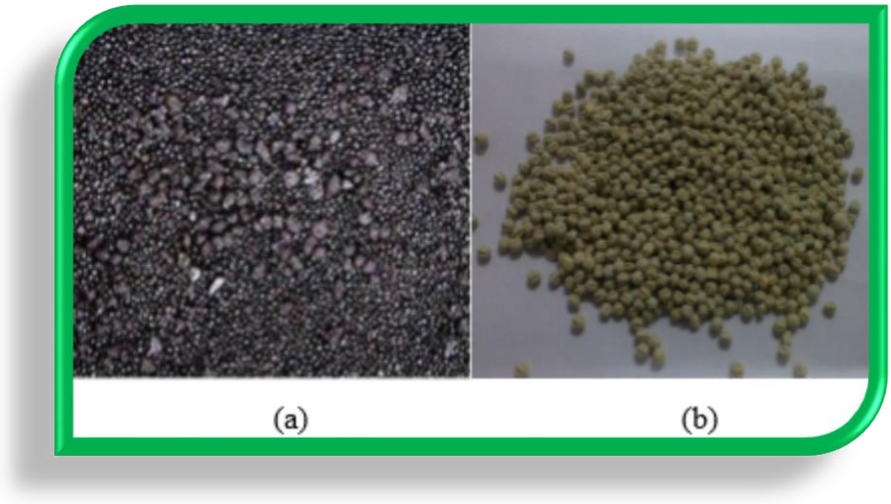


Fig. (2. 4) (a) Coked catalyst, (b) Fresh catalyst

But the nature of the reactions carried out over metallic or acidic catalysts is such that fouling is more likely over metallic or acidic catalysts. In general, fouling involves coke formation, or the deposition of inorganic materials on a catalyst.

The selection of a heterogeneous catalyst for a given action depends not only on the fact that the conversion of feedstock to desired products is high on the chosen catalyst, but also on the fact that the catalyst remains active for long periods under the particular conditions of operationTable (2.4): properties of Major synthetic Zeolites. (Figueiredo, 1981).

Table (2. 3) properties of Major synthetic Zeolites

Zeolite Type	Pore Size Dimensions (A °)	Silica to Alumina Ratio	Application
Zeolite A	4.1	2-5	Detergent manufacturing
Faujasit	7.4	3-6	Catalytic cracking and hydrocracking
ZSM-5	5.2*5.8	30-200	Xylene isomerization, benzene alkylation, catalytic cracking, catalyst methanol conversion

2.6.3 type of the catalyst:

Characterized by ultra-stable Y zeolite as active compound and contains vanadium resisting agent. It has high hydrothermal stability and capable of cracking heavy oil. It is a coke and light oil selective catalyst. The applications: **(Anon., 2013)**

- For processing residue and feeds difficult to be cracked.
- Feeds with high vanadium, sodium, calcium and basic nitrogen.
- For RFCC units required lower slurry yield.
- Higher activity and stability which allows RFCC to be operated at lower C/O ratio.

Chapter (3)
Methodology

The main feed stream crude oil data from Nile blend from Central Petroleum Lap rotary (CPL) and data operation from (CDU-RFCC) Unit.

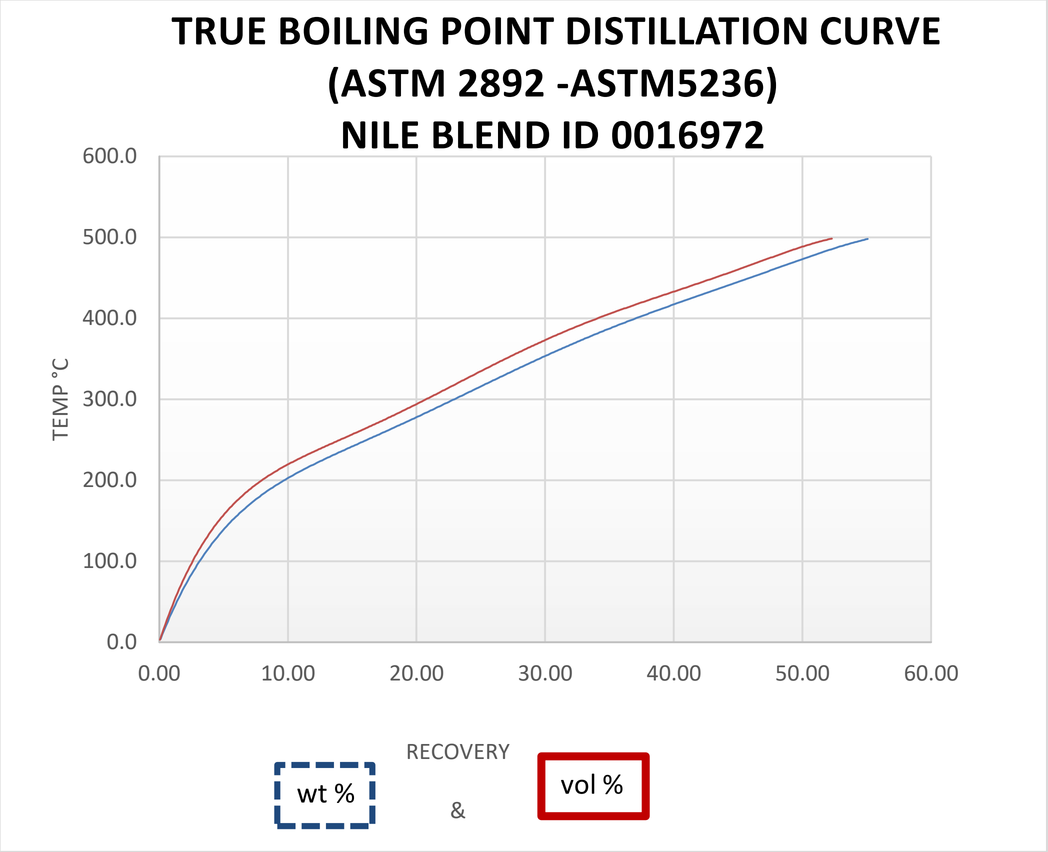


Fig. (3. 1) TBP petroleum assay

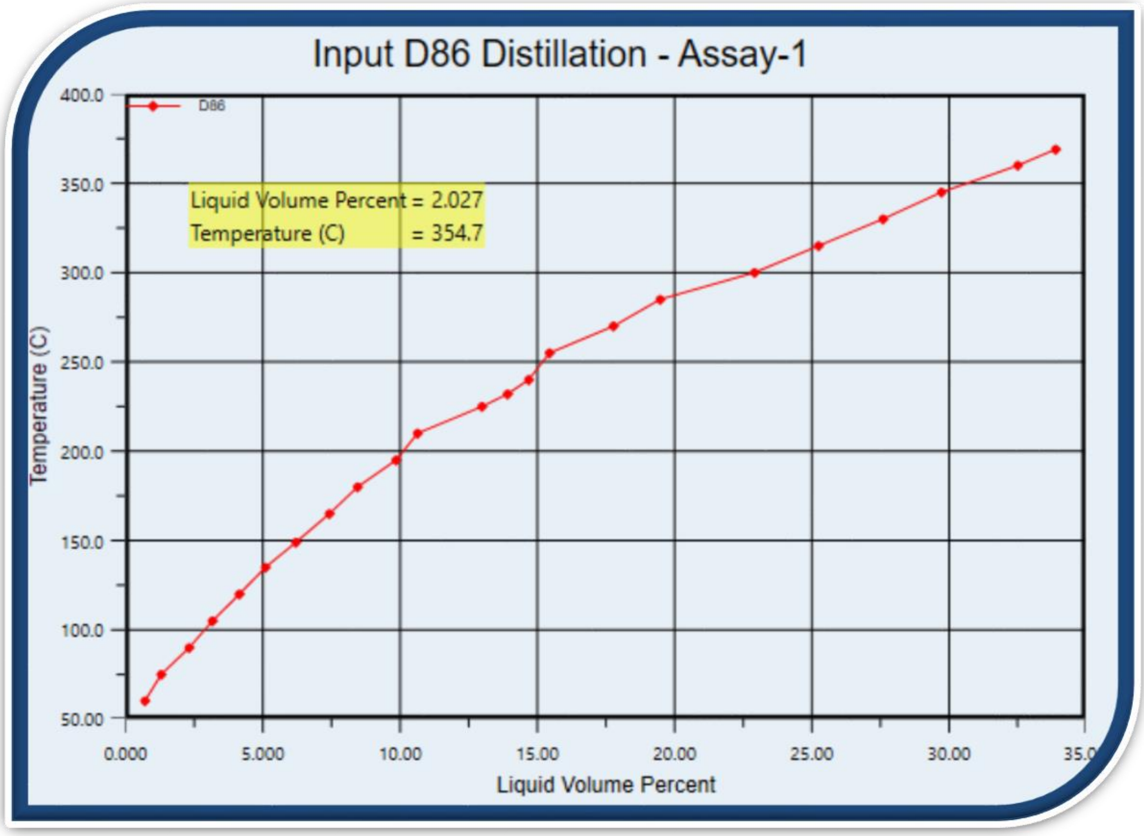


Fig. (3. 2) Input D86 Distillation-Assay-1

3.2. Case study:

The effect of air volume in the regenerator was studied on the reactivation of the catalytic agent which in turn increased the production of gasoline by the unit RFCC. This study was applied to the simulation program.

3.3. Important Reactions Occurring in the FCC Unit:

3.3.1. Thermal cracking:

Before the advent of the catalytic cracking process, thermal cracking was the primary process available to convert low-value feedstocks into lighter products. Refiners still use thermal processes such as delayed coking and tiebreaking for cracking of residual hydrocarbons.

Thermal cracking is a function of temperature and time. The reaction occurs when hydrocarbons in the absence of a catalyst are exposed to high temperatures in the range of (800f-1,200F) (425c-650C). Equation (6.1) shows formation of a free radical when a paraffin molecule is thermally cracked

$$R_1 - CH_2 - CH_2 - R_2 \longrightarrow R_1 - CH_2 + CH_2 - R_2$$

Free radicals are extremely reactive and short-lived. They can undergo alpha-scission, beta scission, and polymerization. (Alpha-scission is a break one carbon away from the free radical; beta-scission, two carbons away.). (Speight, 2007)

Once formed, carbenium ions can form a number of different reactions. The nature and strength of the catalyst acid sites influence the extent to which each of these reactions occur. (Sadeghbeigi, 2012)

The three dominant reactions of carbenium ions are:

1. The cracking of a carbon bond

2. Isomerization.

3. Hydrogen transfer.

3.3.2. Cracking Reaction:

a) Paraffin’s cracking to olefin and smaller paraffin’s.

$$C_{10}H_{22} \rightarrow C_3H_6 + C_3H_{16}$$

B) Olefins cracking to smaller olefins.

$$C_8H_{16} \rightarrow C_5H_{10} + C_3H_6$$

C) Aromatics side chain reactions.

$$Ar - C_{10}H_{21} \rightarrow Ar - C_5H_9 + C_5H_{12}$$

D) Napthenes cracking to olefins and smaller rings

$$\boldsymbol{CYclo - C_{10}H_{20} \rightarrow C_6H_{12} + C_4H_8}$$

3.3.3. Isomerization Reactions.

Isomerization reactions occur frequently in catalytic cracking, infrequently in thermal cracking. In both, breaking of a bond is via beta-scission. However, in catalytic cracking, carbocation's tend to rearrange to form tertiary ions. Tertiary ions are more stable than secondary and primary ions; they shift around and crack to produce branched molecules. In thermal cracking, free radicals yield normal or straight-chain compounds. (Sadeghbeigi, 2012).

- **Some of the advantages of isomerization are as follows:**

1. Higher octane in the gasoline fraction. Isoparaffins in the gasoline boiling range have higher octane than normal paraffin's.
2. High-value chemical and oxygenate feedstock's in the C3/C4 fraction. Isobutylene and isoamylene are used for the production of MTBE and tertiary amyl methyl ether (TAME). MTBE and TAME can be blended into the gasoline to reduce auto emissions.
3. Lower cloud point in the diesel fuel. Isoparaffins in the LCO boiling range improve the cloud point

$$\mathbf{CH_3 - CH_2 - CH_3 - CH_2 - CH_2R \longrightarrow CH_3 - CHCH_3 - CH_2R}$$

3.3.4. Hydrogen Transfer Reactions

Hydrogen transfer is more correctly called hydride transfer. It is a bimolecular reaction in which one reactant is an olefin. Two examples are the reaction of two olefins and the reaction of an olefin and a naphthene.

In the reaction of two olefins, both olefins must be adsorbed on active sites that are close together. One of these olefins becomes a paraffin and the other becomes a cyclo-olefin as hydrogen is moved from one to the other. Cyclo-olefin is now hydrogen transferred with another olefin to yield a paraffin and a cyclodiolefin.

Cyclodiolefin will then rearrange to form an aromatic. The chain ends because aromatics are extremely stable. Hydrogen transfer of olefins converts them to paraffin’s and aromatics. (Sadeghbeigi, 2012)

$$\mathbf{nH_2 \longrightarrow 3C_nH_{2n} + 2 \ + C_nH_{2n} - 6}$$

3.3.5. Other Reactions.

Cracking, isomerization, and hydrogen transfer reactions account for the majority of cat cracking reactions. Other reactions play an important role in unit operation. Two prominent reactions are dehydrogenation and coking. (Sadeghbeigi, 2012)

3.3.6. Dehydrogenation:

Under ideal conditions, i.e. a “clean” feedstock and a catalyst with no metals, cat cracking does not yield any appreciable amount of molecular hydrogen. Therefore, dehydrogenation reactions will proceed only if the catalyst is contaminated with metals such as nickel and vanadium. (Sadeghbeigi, 2012)

3.3.7. Coking:

Cat cracking yields a residue called coke. The chemistry of coke formation is complex and not very well understood. Similar to hydrogen transfer reactions, catalytic coke is a “bimolecular” reaction. It proceeds via carbenium ions or free radicals. In theory, coke yield should increase as the hydrogen transfer rate is increased. It is postulated .that reactions producing unsaturated and multiring aromatics are the principal coke-forming compounds. Unsaturates such as olefins, diolefins , and multiring polycyclic olefins are very reactive and can polymerize to form coke. (Sadeghbeigi, 2012)

Table (3. 1) Some Thermodynamic Data for Idealized Reactions of Importance in Catalytic Cracking

1. Cracking: Paraffin's cracked to olefins and smaller paraffin's Olefins cracked to smaller olefins Aromatic side-chain scission Naphthenes (cycloparaffins) cracked to olefins and smaller ring compounds	$C_{10}H_{22} \rightarrow C_4H_8 + C_6H_{12}$ $C_9H_{18} \rightarrow C_4H_8 + C_5H_{10}$ $ArC_{10}H_{21} \rightarrow ArC_5H_9 + C_5H_{12}$ $CYCLO - C_{10}H_{20} \rightarrow C_6H_{12} + C_4H_8$
2. Isomerization: Olefin bond shift Normal olefin to iso-olefin Normal paraffin to isoparaffin Cyclohexane to cyclopentane	$1 - C_4H_8 \rightarrow trans - 2 - C_4H_8$ $n - C_5H_{10} \rightarrow iso - C_5H_{10}$ $n - C_4H_{10} \rightarrow iso - C_4H_{10}$ $C_6H_{12} + C_5H_9CH_3$
3. Hydrogen transfer: Cycloaromatization	$Naphthene + olefin \rightarrow aromatic + paraffin$ $C_6H_{12} + 3C_5H_{10} \rightarrow C_6H_6 + 3C_5H_{12}$
4.Trans alkylation/alkyl-group transfer	$C_8H_{10} + C_6H_6 \rightarrow 2C_6H_5CH_3$
5. Cyclization of olefins to naphthenic	$C_7H_{14} \rightarrow CH_3 - cyclo - C_6H_{11}$
6. Dehydrogenation	$n - C_8H_{18} \rightarrow C_8H_{16} + H_2$
7. DE alkylation	$iso - C_3H_7 - C_6H_5 \rightarrow C_6H_6 + C_3H_6$
8.Condensation	$Ar - CH = CH_2 + R_1CH = CHR_2 \rightarrow Ar - Ar + 2H$

3.4 Processes description

Reaction and Regeneration Section:

The unit is designed to handle two kinds of feeding systems, hot feeding and cold feeding. Hot feedstock from the atmospheric distillation unit (CDU) and the delayed cocking unit is directly fed to the feedstock surge drum while in cold feeding, the cold feedstock is drawn from unit tank farm using the feedstock pump to the surge drum and then drawn out using the feedstock pump or directly from the unit tank without passing through the surge drum and then heated to 150-200 c in the slurry /feed heat exchanger prior to being fed into the lower section in the riser. Antimony passivation solution is injected into the inlet header of feedstock using the pumps to suppress the action of the heavy metals contaminants and at the same injection point, anti-coke agent is injected.

The feedstock heated is divided into 4 flow valves and inject (atomized) into the riser and converted into fine droplet through 8 nozzle using atomizing steam at two location above each other where only 2 nozzle below and it 6 above. The atomizing steam rate is designed to be 6 to 8% of the feed stock mixes with seceding hot regenerated catalyst from the 2nd regenerator, it vaporizes and the catalyst and oil are accelerated in the upward direction in the cracking reaction take place during the residence time in the riser (2-4 sec) producing the cracked oil vapour which is mix light product as LPG. Gasoline and diesel as well as dry gas, slurry and coke that deposited on the catalyst. (Anon., 2013)

The hot regeneration catalyst enters the riser at the WYE section below the feed injection point and pushed up by left steam ring located at the bottom of the riser then 4 nozzles of left dry gas.

The specially designed feedstock injection nozzles guarantee the maximum conversion of the oil to light-end product, and minimize the coke formation Coker naphtha nozzles were added as 2 below and 2 above the upper feedstock nozzles. Through 2 control valve 4 nozzles. Recycle oil is injected in the riser above the feed injection point to increase the conversion per cycle.

The riser is terminated by 4 quick cyclone separator spent catalyst is separated from the hydrocarbon vapours and dropped down through the dip legs to the stripping section, the dip legs of the cyclone are submerged inside the catalyst bed that is kept in fluidized state by the steam via steam rings . (Anon., 2013)

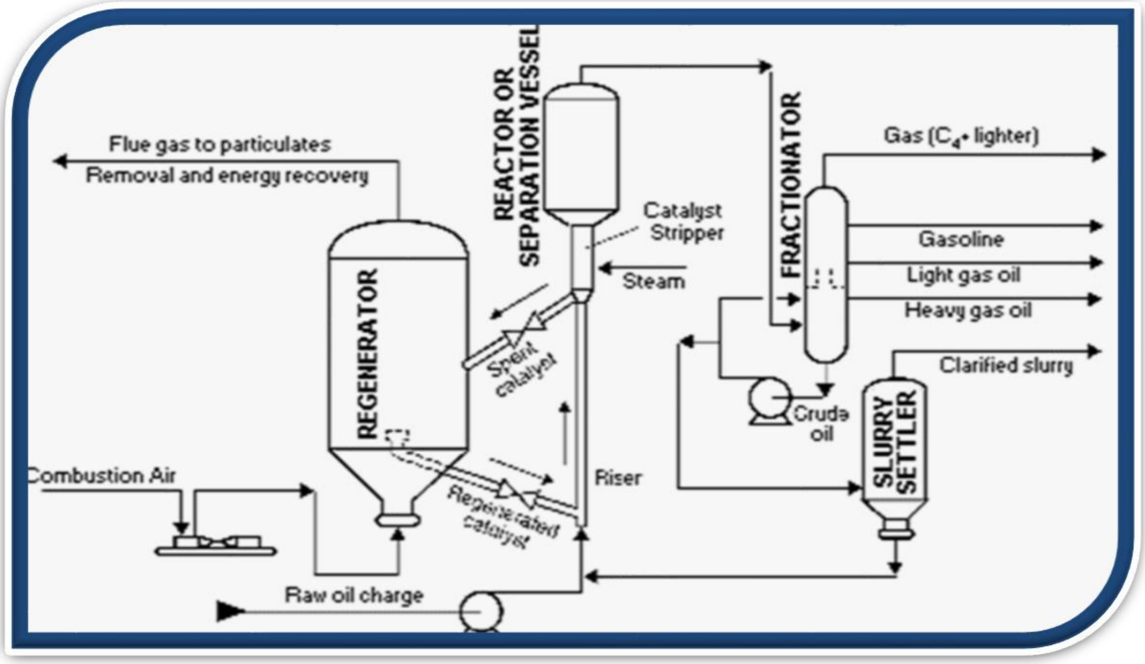


Fig. (3.3) Schematic of the fluid catalytic cracking unit (Speight, 2007)

3.5. ASPEN HYSYS simulator:

Provides an opportunity to check the feasibility of a process, to study and investigate the effect of various operating parameters on various reactions. It is a strong tool for simulation studies and helps in analysing the outcome of a process. HYSYS offers a high degree of flexibility because there are multiple ways to accomplish specific tasks. This flexibility combined with a consistent and logical approach to how these capabilities are delivered makes HYSYS an extremely versatile process simulation tool. The usability of HYSYS is attributed to the following four key aspects of its design:

1-Event Driven operation
2-Modular Operations
3-Multi-flow sheet Architecture
4-Object Oriented Design

3.6. Overall balance:-

3.6.1. Flow sheet wide Mass balance:

Inlet Streams	Mass flow (tonne/hr.)	Outlet Streams	Mass flow (tonne/hr.)
Nile Blend	267.9	naphtha	26.41
Steam	2.145	water	2.102
-	-	Vapour product	380.2

3.6.2. Flow sheet wide energy balance:

Inlet Stream	Energy flow (Btu/hr.)	Outlet Streams	Energy flow (Btu/hr.)

Nile Blend	-4.250e+008	Q2	5.395e+005
q10	1.197e+008	naphtha	-5.246e+007
Steam	-2.575e+007	Q-Condenser	2.084e+007
Q-100	-1.227e+008	water	-3.130e+007
		Vapour product	-4.905e+008

3.7. Material and Energy Balance:

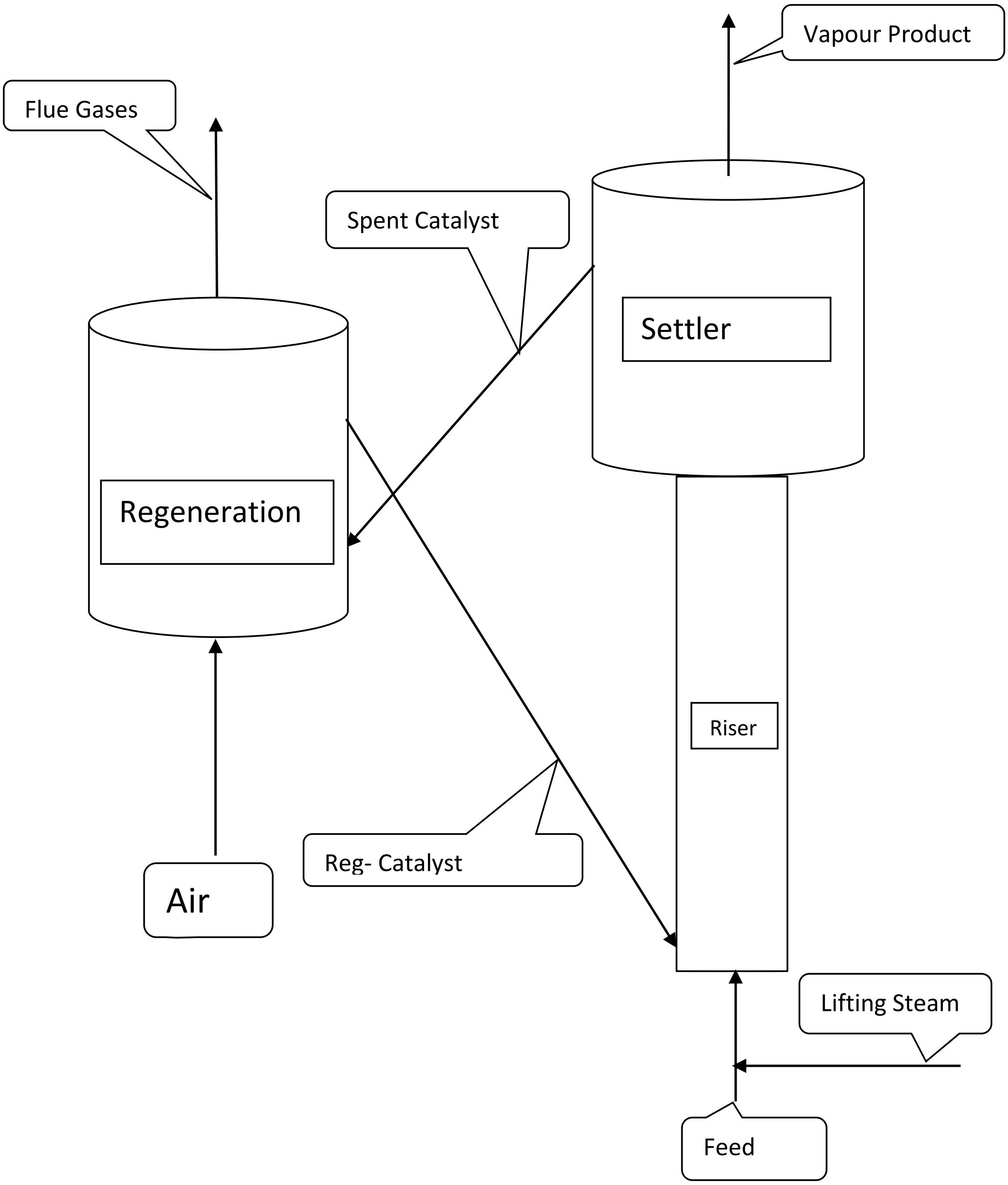
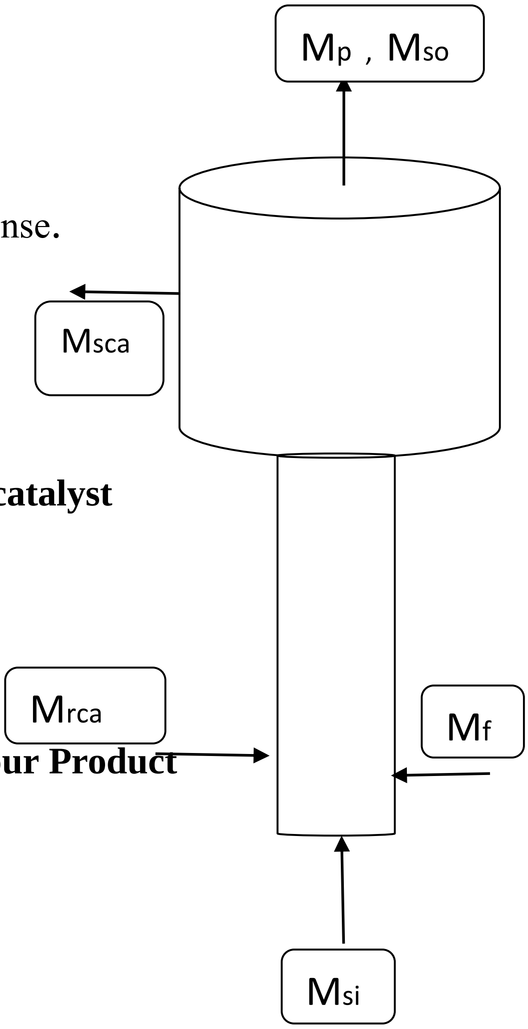


Fig. (3. 3) Input and output streams for reactor and regenerator in FCCU.

- ❖ To material and energy balance around the reactor and the Regenerator can be calculated by defining the input and output streams (Figure 3.2).

3.7.1 Reactor Material Balance:



- Assuming system is steady state.
- Assuming steam inlet doesn’t condense.

- **Mass in**

Feed + steam inlet +Regenerated catalyst

- **Mass out**

Spent catalyst + steam out + Vapour Product

Msca = Mock + Mrca

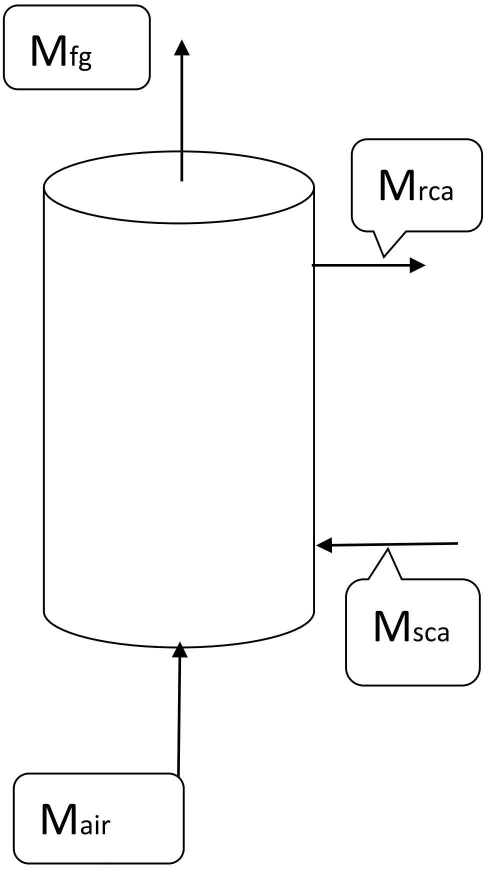
- ✓ Over all material balance around reactor

Mass in = Mass out

Mf + Msi + Mrca = Mock + Meca + Mso + Mp

Mf = Mcok + Mp

3.7.2 REGENERATOR MATERIAL BALANCE:



- **Mass in**

Air blower + Spent catalyst

- **Mass out**

Flue gases + Regenerated catalyst

- ✓ Over all material balance around regeneration

Mair + Msca = Mfg. + Mrca

3.8. Reactor Design:

In Khartoum refinery like in modern RFCC units, total riser reactor is applied, where reactions completely take place in the riser, the reactor consists of the following parts.

3.8.1- Riser:

It is a long vertical straight pipe where the desired reactions take place, quick separation of the catalyst from hydrocarbons vapor at the end of the riser important to give good result.

3.8.2. Riser Calculation:

First we must to calculate the volumetric flow rate of feed,

☒ ***Volumetric flow rate of feed (V):***

The volumetric flow rate can be calculated by the following equation:

$$\boldsymbol{V = Mcat[\frac{1}{\rho b} - \frac{1}{\rho c}]} \qquad \text{(Anon., 2013)}$$

The density of the bulk differs between the top and bottom of the riser, so the volumetric flow rate must be calculated at the both side.

i. ***The volumetric flow rate at out let of the riser (V◦):***

☒ At the outlet of the riser we find:

M cat= 1647498 [*material balance]*

ρ_c= 780 kg/m³

ρ_b=16 kg/ m³ (Anon., 2013)

V= 1647498[1/16 -1/780]=10085.44 m³/hr. =28 m³/s

ii. ***the volumetric flow rate at the inlet of the riser (Vi):***

☒ At the inlet we have:

M cat= 1647498 **[** ***material balance]***

$\boldsymbol{\rho_c}$= 780 kg/m³

$\boldsymbol{\rho_b}$=160 kg/ m³ (Anon., 2013)

V=1647498[1/160 -1/780] =8187.69m³/hr.

= **2.27 m³/s.**

⇒ **The average flow rate can be determine by the following equation:-**

$$\boldsymbol{Vav} = \frac{\mathbf{Vo - Vi}}{\mathbf{ln}(\frac{\mathbf{Vo}}{\mathbf{Vi}})}$$

☒ **Where:-**

Vav: the average volumetric flow rate of feed in the riser,

VO: the volumetric flow rate at the out let of the riser which are calculated previous = 28 m³/s,

VI : the volumetric flow rate at the inlet of the riser which also calculated previous = 2.27 m³/s.

$$\boldsymbol{Vac} = \frac{28 - 2.27}{\ln(\frac{28}{2.27})} = \mathbf{10.24\ m^3/s}$$

The linear velocity through the riser:

The linear velocity of the bottom of the riser should be not less than **(4.5 m/s),** and at the top of the riser should be not less than **(18 m/s)** , **so the average velocity must be calculated as follows**:

$$\boldsymbol{Uav} = \frac{\mathbf{Uo - Ui}}{\mathbf{ln}(\frac{\mathbf{Uo}}{\mathbf{Ui}})}$$

$$Uav = \frac{18 - 4.5}{\ln(\frac{18}{4.5})} = 9.7m/s$$

☒ ***Riser Diameter:***

The average volumetric flow rate

= cross section area * average linear velocity

$$\boldsymbol{Vav = A * Uav}$$

$$\boldsymbol{A = Vav/Uav}$$

$$\boldsymbol{A = \frac{10.24}{9.7} = 1.05m^2}$$

☒ **So riser diameter can be calculated from the following equation:**

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

$$D^2 = 4 * A/\pi$$

$$\boldsymbol{D = \frac{\sqrt{4*1.05}}{3.14} = 1.16m}$$

☒ ***Riser Height:***

☒ Riser height can be calculated by the following equation:

Height (H) = average linear velocity (U_{av}) * residence time (τ)

$$\boldsymbol{H = Uav * \tau}$$

H : riser height

$\boldsymbol{U_{av}}$: Average linear velocity which calculated previous =9.7 m/s,

τ : Residence time in the riser = 5 s (Anon., 2013)

$$\boldsymbol{H = 9.\frac{7m}{s} * 5s = 48.5m}$$

☒ ***Pressure Drop through the Riser:***

Pressure drop can be calculated by the following equation:

$$\boldsymbol{\Delta P = 1.3[\frac{U_g^2\,\rho_g}{2 * g_c} + W * \frac{U_P}{g_c}]}$$

Calculation of the solid flux (W):

☒ **Solid flux (W) can be calculated as follows:**

$$\boldsymbol{W_S = (mass\ flow\ of\ the\ catalyst) / (cross\ secation\ areaof\ riser)}$$

☒ **But,**

Mass flow of the catalyst = 1647498 kg/hr.

Cross section area of the riser = 1.16 m² **[*calculated previously*]**

$$W_S = \frac{\boldsymbol{1647498}}{\boldsymbol{(3600)\left(\frac{\pi}{4}\right)(1.16^2)}} = \boldsymbol{433 kg/m^2.s}$$

☒ **And we have,**

U_g = U_p = 12.48 m/s (Anon., 2013)

ρ_g**= 16 kg/m³,**

g_c **= 1 kg.m/s².N**

☒ **Now we can substitute these values in the main equation:**

$$\Delta P = [\frac{(9.7)^2 * (16)}{2*1} + \frac{(433)(9.7)}{1}$$ = **6438.7 N/m =0.00644Map**

3.8.3- Settler:

It is a vessel used to move the catalyst to the first regenerator

☒ ***Settler Calculation:–***

Settler diameter:

At first settler volume must be calculated, this can be done as follows:

$$V = \frac{\pi}{4(D_o^2 - D_i^2)} * L$$ (Anon., 2013)

It is found that the maximum level of the catalyst used to keep a good stripping is (L= 9)

Volume of settler can be calculated by other equation:

$$V = W_{Ca}/\rho$$

☒ **Where:**

W_{Ca}: Weight of catalyst,

ρ: catalyst density in settler = 700 kg/m³ (Anon., 2013)

The weight of catalyst = catalyst circulation * residence time of catalyst

☒ ***<u>But,</u>***

- Catalyst circulation = 1671123 kg/hr. [Material *balance*]
- Residence time = 4.5 min (Anon., 2013)

<u>So,</u>

$$\mathbf{Wca} = \frac{\mathbf{1671123}}{\mathbf{60}}$$

$$= \mathbf{12\,5334\ \ kg}$$

<u>So,</u>

$$\mathbf{V} = \frac{\mathbf{125334}}{\mathbf{700}}$$

$$= \mathbf{179}\,\boldsymbol{m^3}.$$

D_i (**riser diameter**) = 1.16 m **[*calculated previously*]**

☒ **So, these values can be substituted in the main equation:**

$$\boldsymbol{V = \frac{\pi}{4}\left(D_0^2 - D_i^2\right) * L}$$

$$\mathbf{179} = \boldsymbol{\frac{\pi}{4}}\left(\boldsymbol{D_0^2} - (1.16^2)\right) * \boldsymbol{L}$$

$$D_O^2 = \frac{4*(174+\frac{\pi}{4}(1.16)^2*9)}{\pi*9}$$

$$D_O = \sqrt{\frac{4*(174+\frac{\pi}{4}(1.16)^2*9)}{\pi*9}}$$

= 5.2 m

3.9. Regenerator Design:

3.9.1. The First Regenerator Design:

❖ ***Introduction:-***

Catalyst leaving the reaction section contains a carbonadoes deposit on it surface .the carbon deposited on the catalyst surface occupies active catalyst it deposit along with other forms of coke, is burned in the regenerator this restores the catalyst activity and provides the heat necessary for the cracking reaction.

Regenerators are large vessels where the catalyst and air are mixed efficient operation.

RFCC regenerators are generally equipped with two stage cyclones. In these systems, the outlet duct of the first stage cyclone is coupled directly to the inlet duct of the second stage cyclone system. Most RFCC unit regenerators employ 4-16 parallel sets of primary and secondary cyclones.

☒ ***The dense phase diameter calculation:-***

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

$$V = W_{Ca}/\rho$$

$W = M_{cat} * t$ **[*Catalytic cracking of heavy petroleum]***

Where:-

m_{cat}=the mass flow rate of the catalyst (Kg)

t = the optimum residence time for the catalyst in the 1st regenerator lays between (5-10 min), 5 min is selected.

m_{cat}= 1671123 kg/hr. **(Material balance)**

t =5 min **[*Petroleum Refining*]**

ρ = 780 kg/m³ **[*refinery manual*]**

$$\boldsymbol{W = 1671123 * \frac{5}{60} = 139260.25} \textbf{ kg}$$

$$\boldsymbol{V = \frac{W}{\rho} = \frac{139260.25}{780} = 178.54 m^3}$$

⇒ L = 11m

$$D = \sqrt{\frac{4V}{\pi * L}} = \sqrt{\frac{4*178.54}{\pi*11}} = 4.5\text{m}$$

The dilute phase diameter calculations:

Table (3.2) Flue gas composition [KRC]

Composition	*Wt. %*	*Kg/hr.*	*M*	*Kmol/hr.*
CO	**3.611**	**8530.987**	**28**	**304.678**
CO2	**10.942**	**25850.462**	**44**	**587.511**
N2	**71.658**	**169292.024**	**28**	**6046.144**
O2	**0.173**	**408.7125**	**32**	**12.7723**
H2	**13.616**	**32167.8**	**18**	**1787.1**
Total	***100***	***236250***	**-**	***8738.2***

$$\text{The dilute phase area} = (\text{Volume of the flue gas})/(\text{Liner velocity})$$

$$[\textit{Catalytic cracking}]$$

The volume of the gas V1:

From ideal gas law:

$$V1 = \frac{P_0 V_0 n_1 T_1}{P_1 n_0 T_0}$$

Where:

P0 = 1.01325 N/m² **P1**- Dilute phase pressure = 0.253 MPa
V0 = 22.4 m³ **T1**- Dilute phase temperature = 903 ºK

n0 = 1 kmol **n1**- 8738.2 kmol/hr. **T0** = 273 ºK

$$V = \frac{1.01325 * 22.4 * 8738.2 * 903}{0.253 * 10^6 * 1 * 273}$$

$$= \mathbf{259293.2 m^3/hr} = \mathbf{72.03 m^3/hr}$$

$$A = \frac{Volume}{Linear\ velocity}$$

Linear velocity = 1.9 m/s [*Catalytic cracking*]

$$A = \frac{72.03}{1.9} = \quad 37.90 m^2$$

$$D = \sqrt{\frac{4A}{\pi}} = 6.9\text{m}$$

The dilute phase diameter = 6.9 m

☒ *The thickness of cylinder wall:*

$$\frac{\boldsymbol{P_i D_i}}{\boldsymbol{2f * j - P_i}} + \boldsymbol{C}$$ [*Chem. En. V6*]

Where:

$\boldsymbol{t}$: Thickness (m);

Di: internal diameter = 6.9 m;

F: stress of material = 90N/mm²; from appendix

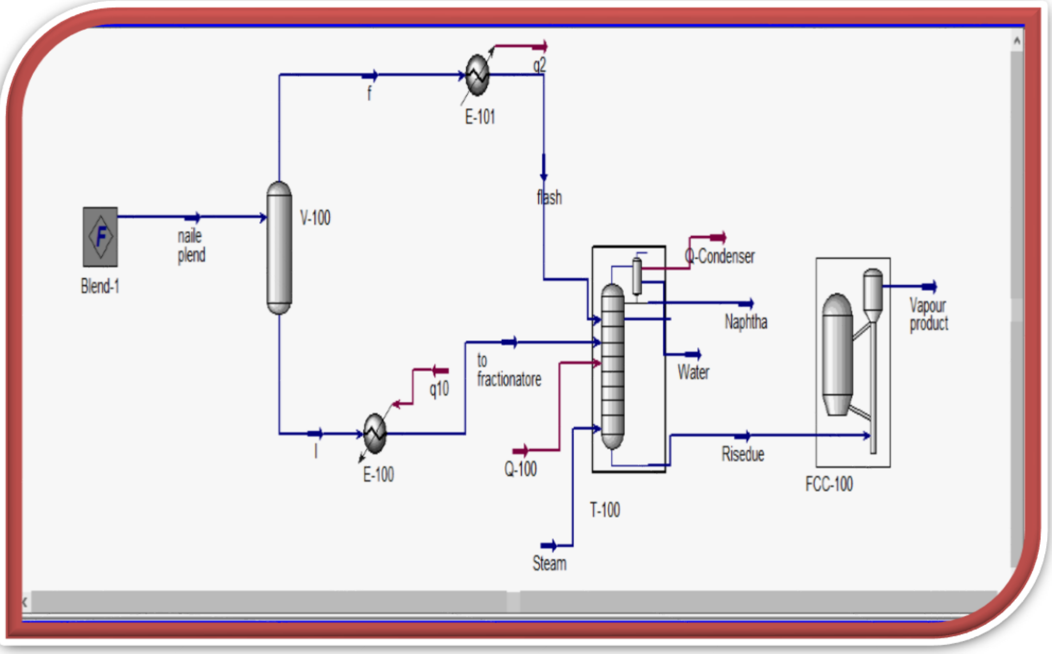
J: welded joint factor;

Pi: internal pressure = 0.253 MPa;

$$\boldsymbol{t} = \frac{\mathbf{0.253 * 6.9}}{\mathbf{2 * 90 - 0.253}} = 0.09712m = 9.712cm$$

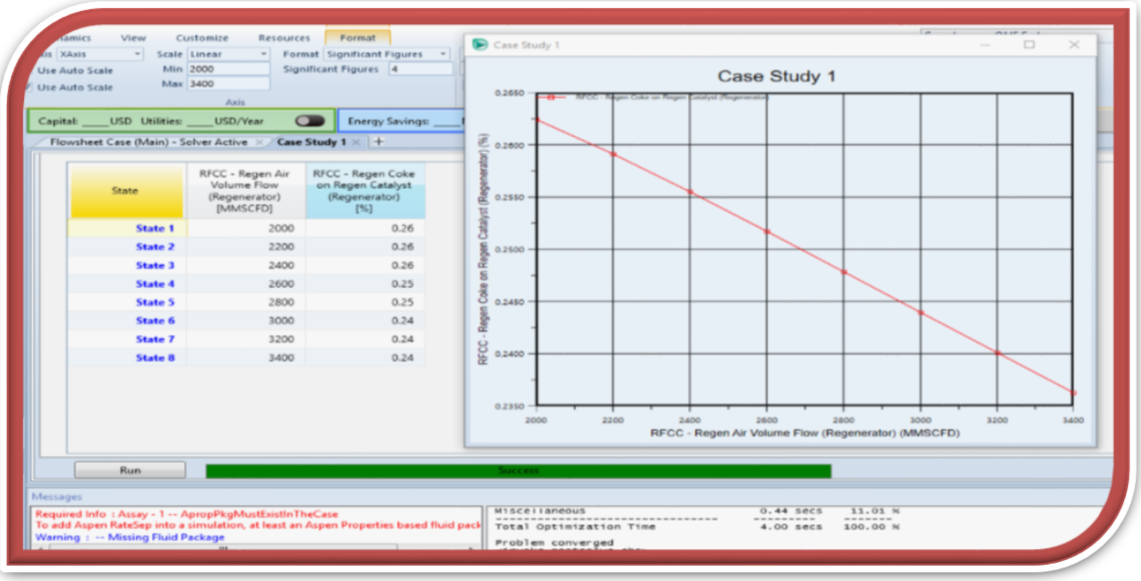
Chapter (4)

4.1. Result and discussion:



Fig(4.1) flow sheet simulation

Case (1):-



is XAxis
Scale Linear
Format Significant Figures
Use Auto Scale
Min 2000
Significant Figures 4
Use Auto Scale
Max 3400

Capital: ____USD Utilities: ____USD/Year

Energy Savings: ____

Flowsheet Case (Main) - Solver Active

Case Study 1

State	RFCC - Regen Air Volume Flow (Regenerator) [MMSCFD]	RFCC - Regen Coke on Regen Catalyst (Regenerator) [%]
State 1	2000	0.26
State 2	2200	0.26
State 3	2400	0.26
State 4	2600	0.25
State 5	2800	0.25
State 6	3000	0.24
State 7	3200	0.24
State 8	3400	0.24

Success

Messages

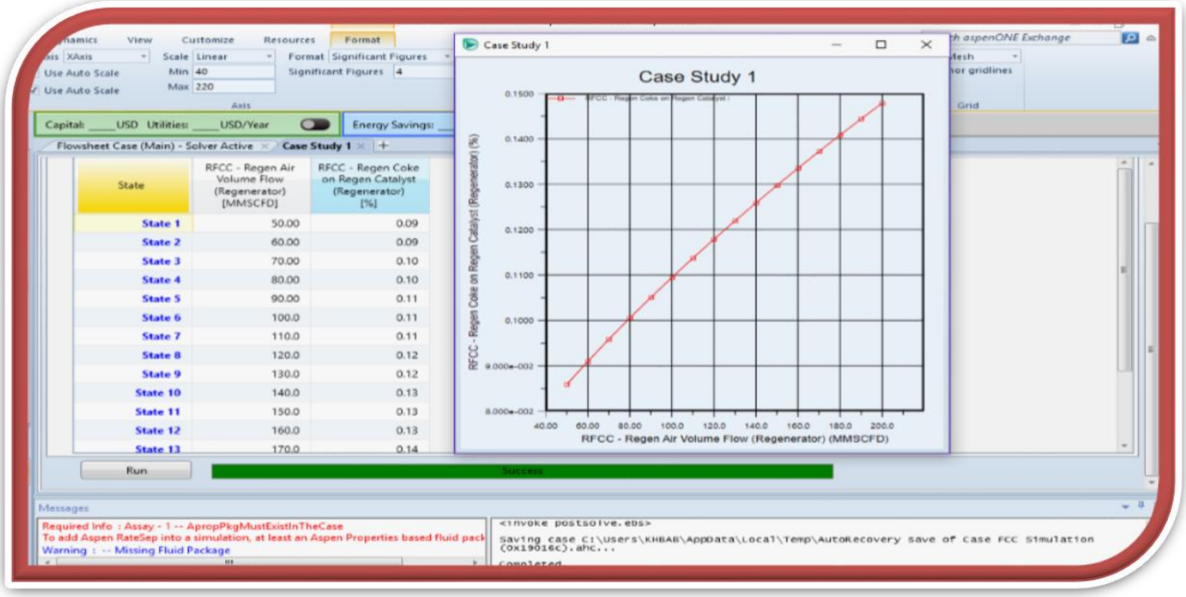
```
Required Info : Assay - 1 -- AspropPkgMustExistInTheCase
To add Aspen RateSep into a simulation, at least an Aspen Properties based fluid pack
Warning : -- Missing Fluid Package
```

```
Miscellaneous                              0.44 secs    11.01 %
-------------------------------------      ---------    -------
Total Optimization Time                    4.00 secs   100.00 %

Problem converged
```

Discussion OF Case (1):-

From this case that obtain of ASPEN HYSYS simulation appearing the amount of carbon removed in regeneration catalyst increases due to amount of air combustion increase.

Case (2):-



State	RFCC - Regen Air Volume Flow (Regenerator) [MMSCFD]	RFCC - Regen Coke on Regen Catalyst (Regenerator) [%]
State 1	50.00	0.09
State 2	60.00	0.09
State 3	70.00	0.10
State 4	80.00	0.10
State 5	90.00	0.11
State 6	100.0	0.11
State 7	110.0	0.11
State 8	120.0	0.12
State 9	130.0	0.12
State 10	140.0	0.13
State 11	150.0	0.13
State 12	160.0	0.13
State 13	170.0	0.14

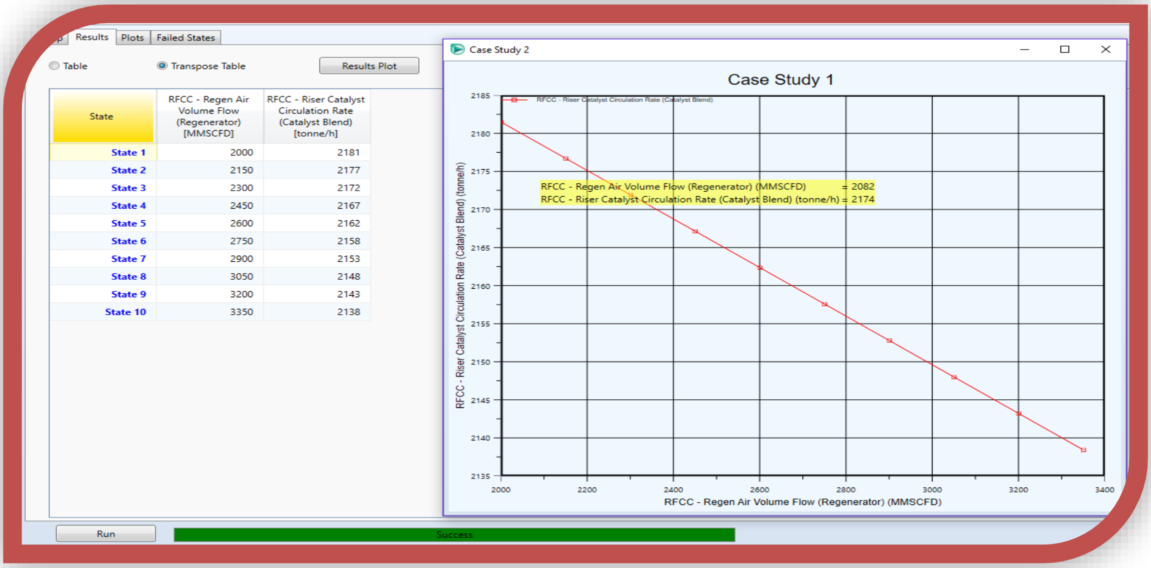
```
Required Info : Assay - 1 -- AspropPkgMustExistInTheCase
To add Aspen RateSep into a simulation, at least an Aspen Properties based fluid pack
Warning : -- Missing Fluid Package
```

```
<invoke postsolve.ebs>

Saving case C:\Users\KHBAB\AppData\Local\Temp\AutoRecovery save of case FCC Simulation
(0x19016c).ahc...

Completed
```

Discussion OF Case (2):-

While the small amount of air combustion the amount of carbon is increase.

Case (3):-



State	RFCC - Regen Air Volume Flow (Regenerator) [MMSCFD]	RFCC - Riser Catalyst Circulation Rate (Catalyst Blend) [tonne/h]
State 1	2000	2181
State 2	2150	2177
State 3	2300	2172
State 4	2450	2167
State 5	2600	2162
State 6	2750	2158
State 7	2900	2153
State 8	3050	2148
State 9	3200	2143
State 10	3350	2138

RFCC - Regen Air Volume Flow (Regenerator) (MMSCFD) = 2082
RFCC - Riser Catalyst Circulation Rate (Catalyst Blend) (tonne/h) = 2174

Discussion OF Case (3):-

When amount of air combustion is increase catalyst circulation rate decrease.

Design riser Discussion

Catalyst circulation rate= 2.115e+006 Kg/h r [HYSYS Results]

Parameter design	Design in [KRC]	Calculation design
Average volume flow	$10.24\ m^3/s$	**13.24 m3 /s**
Area	$1.05m^2$	**1.36 m2**
Diameter	$1.16m$	**0.839 m**
high	$48.5m$	**66.2 m**

Chapter (5)

CONCLUSION & RECOMMENDATION

5.1 CONCLUSION:

In this project we have been study the effect of amount of air in regenerative catalyst in fluid catalytic cracking unit in Khartoum Refinery Company and we have been found the process of catalyst regenerative is effected by amount of air then. And has shown the optimum amount to achieve carbon percentage of less than 0.05% on regeneration catalyst is **3400** Nm^3/min .and design riser calculation Diameter is **0.839 m** and high is **66.2m**

5.2 Recommendation:

For further study of regeneration catalyst technology we recommended by:

1- Pretreatment to the feedstock of unit to reduce the amount of metals such as sodium.
2- check of carbon percentage on regeneration catalyst must be daily

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