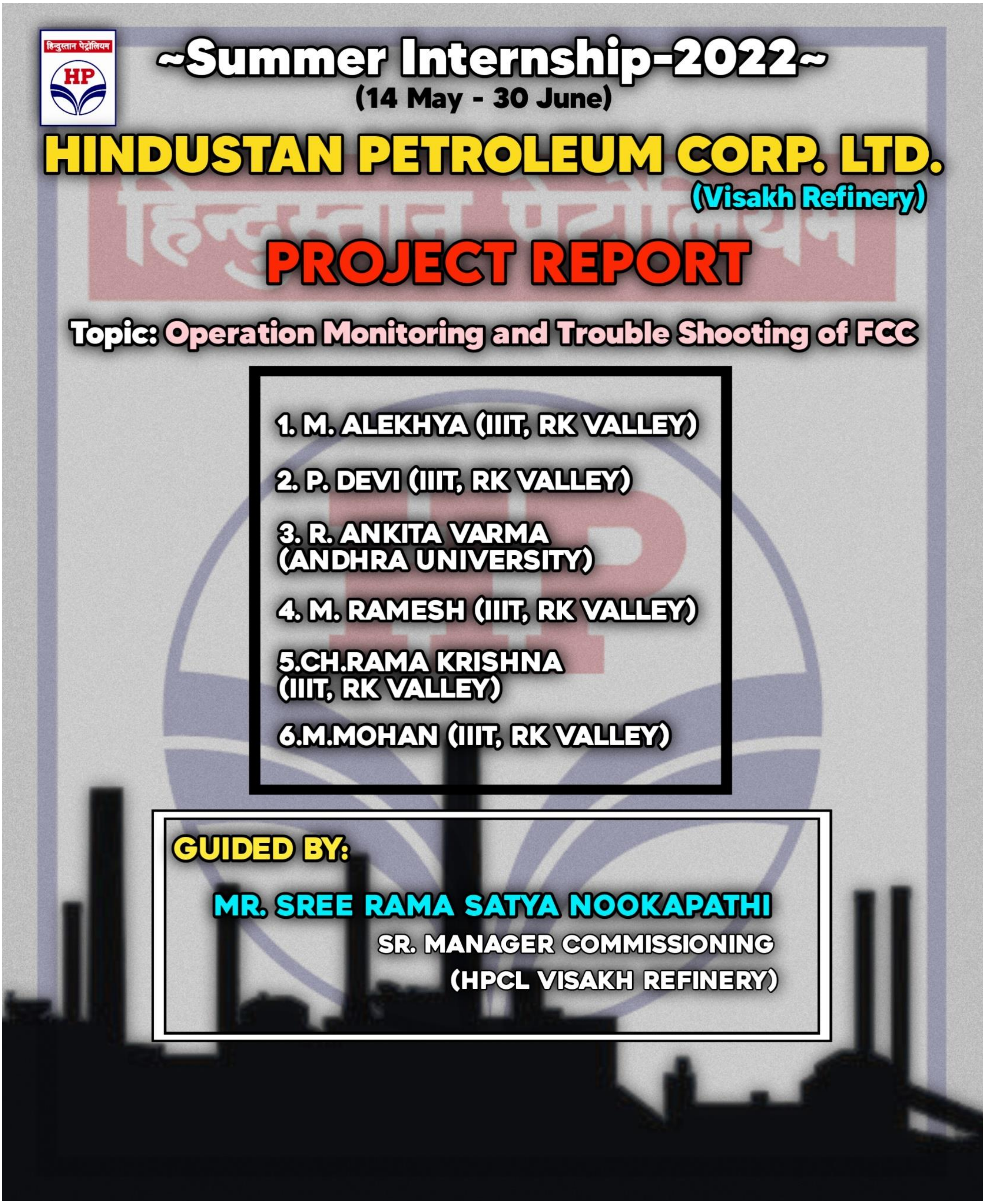

~Summer Internship-2022~

(14 May - 30 June)

HINDUSTAN PETROLEUM CORP. LTD.

(Visakh Refinery)

PROJECT REPORT

Topic: Operation Monitoring and Trouble Shooting of FCC

1. M. ALEKHYA (IIIT, RK VALLEY)
2. P. DEVI (IIIT, RK VALLEY)
3. R. ANKITA VARMA (ANDHRA UNIVERSITY)
4. M. RAMESH (IIIT, RK VALLEY)
5. CH.RAMA KRISHNA (IIIT, RK VALLEY)
6. M.MOHAN (IIIT, RK VALLEY)

GUIDED BY:

MR. SREE RAMA SATYA NOOKAPATHI
SR. MANAGER COMMISSIONING
(HPCL VISAKH REFINERY)

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CHAPTER-1

ACKNOWLEDGEMENT

A Great opportunity from over a period can be successful with the advices and support from HPCL.I take this opportunity to express my gratitude and appreciation to all those who encouraged to complete this project First and foremost, I would like to express my gratitude towards Hindustan petroleum corp.Ltd (HPCL) for providing me as magnificent opportunity in the form of this internship to learn and get the experience. At very first I would like to express my profound and sincere thanks to my project mentor Mr.Sree Rama Satya (senior manager- commissioning) who guided through the project, And Special thanks to HR officer Nithish Bharadwaj,who took out this from his busy schedule and provided valuable guidance and support.

1.1. HINDUSTAN PERTOLEUM CORP .LTD:

HPCL owns and operates Refineries in Mumbai and Visakhapatnam with a capacity of 9.5 Million Metric Tonnes Per Annum (MMTPA) and 8.3 MMTPA, respectively. HPCL also owns India’s largest Lube Refinery in Mumbai, producing lube oil base stocks with a capacity of 428 Thousand Metric Tonnes Per Annum (TMTPA). Both Mumbai and Visakh Refineries have been upgraded to produce BS-VI compliant transportation fuels.

HPCL holds a 48.99% equity stake in the JV company, HPCL-Mittal Energy Limited (HMEL), which operates an 11.3 MMTPA capacity Refinery at Bhatinda in Punjab. It also owns a 16.96% equity stake in the 15 MMTPA Mangalore Refinery and Petrochemicals Ltd. (MRPL). A new 9 MMTPA Greenfield Refinery-cum-Petrochemical complex is being set up at Pachpadra in the Barmer district of Rajasthan through a joint venture company, HPCL Rajasthan Refinery Limited (HRRL) with equity stake of 74%.

HPCL has its Research & development centre named HP Green R&D centre in Bengaluru.The centre provides adavanced technical support to the Refineries and marketing SBUs for opertaional improvement ,absorb new technologies,develop innovative & path breaking technolgies,licence technologies and became a knowledge hub. HPCL is commited to conducting business with an objective of presenting the enviroment sustainble development ,being a safe work place and enrichment of the quality of life of employees.customers and the commuity

STOARGE TANKS AND PIPE LINE SYSTEMS IN VISAKH REFIENERY

1.2. VISAKH REFINERY:

Visakha refinery was commissioned by Caltex Oil Refining (India) in 1957, with an installed capacity of 0.675 million_tonne per year.The facility was acquired by the Government of India in 1976 and amalgamated with HPCL in 1978 by the CORIL-HPCL Amalgamation Order, 1978.The refinery's capacity was increased to 4.5 million tonne per year under the first major expansion project in 1985, known as Visakha Refinery Expansion Project-I (VREP-I).Its refining capacity was increased to 7.5 million tonne per year under VREP-II in 1999, and again to 8.3million tonne per year in 2010.

1.3. V PRODUCT PROFILE:

LIGHT ENDS	**MIDDLE DISTALLATES**	**HEAVY ENDS**
Propylene	kerosene	Fuel
LPG	ATF	LSHS
MS(BS-VI)	Diesel(BS-VI)	Bitumen
Naphtha	HFHSD	
	LDO	
	JBO	

1.4. PROCESS UNITS INSTALLED IN VISAKH REFINERY

The HPCL visakh refinery is an integrated refinery facility spread over 723 acres.

1.3 crude Distillation units (CDUs),

CDU-1:1.8 MMTPA

CDU-2: 3.1MMTPA

CDU-3 :3.4MMTPA

2.2 Fludized catalytic cracking units,

FCCU-1(Exxon):1MMTPA

FCCU-2(UOP): 0.96 MMTPA

3.Vis -Breaker unit:1MMTPA

4.Bitumen Blowing unit: 225 MMTPA

5.propylene Recovery unit:65,000 MMTPA

6.LPG &ATF units

7.Diesel Hydro Desulphuration unit(DHDS) for HSD treatment :2.4 MMTPA (E-III HSD Production)

8.Diesel Hydro treating (DHT) unit for HSD treament : 1.32 MMTPA (E-IV HSD PRODUCTION)

1.5. UTILITIES IN VISAKH REFIENERY

- Cooling water system
- Air compressors
- Boilers for steam generation
- nitrogen system
- Flare system

1.6. VISAKH REFINERY MODERNISATION PROJECT [VRMP]:

A new CDU/VDU-IV with a processing capacity of 9Mtpa will replace one of the three existing CDUs, while the other two CDUs continue to process 3Mtpa each. A full conversion hydrocracker unit of 2.8Mtpa capacity will be installed to process the excess gas oil after saturating the existing

FCCs. All the sour gas oil is processed in once through hydro-cracker unit (OHCU) and the FCC feed comprises OHCU bottoms and sweet vacuum gas oil (VGO).

The VGO hydro-treater is considered only for preparing feed for the new FCC-PC unit whereas the existing FCCs will continue to process the sour VGO.The modernisation project also includes the installation of new residue upgradation units to process the entire vacuum residue for the 15Mtpa capacity refinery. A new solvent deasphalting unit of 2.55Mtpa capacity will be installed to process the balance vacuum residue after saturating the existing VBU and BBU.An 200ktpa additional alkylation unit, a 2.8Mtpa slurry hydrocracker, and a 128tpd propylene recovery unit (PRU) will also be installed as part of the modernisation project. Auxiliary units will include the sulphur and amine recovery units, as well as the LPG treating units.

IMAGE: HUGE VACCUM COLUMN AT VRMP PROJECT

- CDU unit -IV
- Hydrocracking unit (FCHCU)
- Hydrogen Generation unit
- sulphur recovery unit
- CPP
- Utilities(CNU,cooling Towers
- Effluent Treatment plant (ETP Block)
- Flare system

CHAPTER -2

OVERVIEW OF THE COMPANY

Hindustan Petroleum Corporation Limited(HPCL) is a subsidiary of Oil and Natural Gas Corporation(ONGC) which is under the ownership of the Ministry of Petroleum and Natural Gas of the Government of India with its headquarters in Mumbai, Maharashtra. It has a 25% market share in India among public sector undertakings (PSUs) coupled with a strong marketing infrastructure. Its parent company is ONGC which owns a majority stake in the company. The company is ranked 367th on the Fortune Global 500 list of the world's biggest corporations as of 2016. On 24th October 2019, the compa became a Maharatna PSU.

HPCL was incorporated in 1974 after the takeover and merger of erstwhile Esso Standard and lube India Limited by the Esso (Acquisition of Undertakings in India) Act 1974. Caltex Oil Refining (India) Ltd. (CORIL) was taken by the Government of India in 1976 and merged with HPCL in 1978 by the CORIL- HPCL Amalgamation Order, 1978. Kosan Gas Company was merged with HPCL in 1979 by the Kosangas Company Acquisition Act,1979.

On 19th July 2017, the Government of India announced the acquisition of the Hindustan Petroleum Corporation by Oil and Natural Gas Corporation. On 1 November 2017, the Union Cabinet approved ONGC for acquiring a majority 51.11% stake HPCL (Hindustan Petroleum Corporation Limited). On 30th January 2018, Oil & Natural Gas Corporation acquired the entire 51.11% stake of Hindustan Petroleum Corporation, thus becoming the promoter of the company.

HPCL operates two major refineries producing a wide variety of petroleum fuels and specialties, one in Mumbai (Wes Coast) with 7.5 million tonnes per year capacity and the other in Visakhapatnam (East Coast) with a capacity of 8.3 million tonnes per year. HPCL holds an equity stake of 16.95% in Mangalore Refinery and Petrochemicals Limited(MRPL), a state-of-the-art refinery at Mangalore with a capacity of 9 million tonnes per year.

Another refinery of 9 million tonnes per year (set up in Bathinda Punjab by HMEL, a joint venture with Mittal Energy Investments Pte. Ltd). HPCL has signed a memorandum of understanding with a Government of Rajasthan for setting up a refinery near Barmer. It would be operated under a joint venture company (JVC) called HPCL-Rajasthan Refinery Limited.

HPCL also owns and operates the largest lubricant refinery in India producing lube base oils of international standards, with a capacity of 335 TMT. This refinery accounts for over 40% of India’s total lube base oil production. Presently HPCL produces over 300+ grades of lubes, specialties, and greases. The marketing network of HPCL consists of 21 zonal offices in major cities and 128 regional offices facilitated by a supply and distribution infrastructure comprising terminals, aviation service facilities, liquefied petroleum gas (LPG) bottling plants, lube filling plants, inland relay depots, retail outlets (petrol pumps) and LPG and lube distributorships. HPCL has a state-of-the-art information technology infrastructure to support its core business. The data center is at Hi-tech city in Hyderabad.

Vision:

"To be a world-class energy company known for caring and delighting the customers with high-quality products, innovative services across domestic & international markets with aggressive growth and delivering superior financial performance.

The company will be a model of excellence in meeting social commitment, environment, health and safety norms and in employee welfare & relations".

Mission:

"HPCL along with its joint ventures will be a fully integrated company in the hydrocarbons sector of exploration and production, refining and marketing; focusing on enhancement of productivity, quality & profitability, caring for customers and employees, caring for environment protection and cultural heritage. It will also attain scale dimensions by diversifying into other energy-related fields and by taking up transnational operations".

VISAKH REFINERY:

Visakh Refinery is one of the two oil refineries of HPCL in India, the other being Mumbai Refinery. This was one of the first major industries of Visakhapatnam and first oil refinery on the East Coast. After the nationalization, HPCL has transformed itself into a mega Public Sector Undertaking and it is second largest integrated oil company in India. It was commissioned by Caltex Oil Refining (India) in 1957,with an installed capacity of 0.675 million tonne per year. The facility was acquired by the Government of India in 1976 and amalgamated with HPCL in 1978 by the CORIL-HPCL Amalgamation Order,1978.

The Refinery's Capacity was increased to 4.5 million tonne per year under the first major expansion project in 1985, known as Visakh Refinery Expansion Project-I (VREP-I). Its Refining capacity was increased to 7.5 million tonne per year under VREP-II in 1999, and again to 8.3 million tonne per year in 2010.

Visakh Refinery, till 1985, used to receive indigenous and imported crude through tankers cargo up to 30 TMT from the Oil Wharf Jetties located in Visakhapatnam Port Trust (VPT). Offshore tanker Terminal (OSTT) was commissioned by VPT for crude unloading through Suez max tanker. The draft limitations at Jetty were restricting bigger size vessels. To overcome this limitation, Visakh Refinery constructed Single Point Mooring (SPM) for discharging of Very Large Crude Carrier (VLCC).

The products currently manufactured in Refinery are LPG, Propylene, Naphtha, MS, ATF, SKO,HSD, LDO, LSHS, IFO, Bitumen, Sulfur, VLSFO. These products are evacuated to the marketing terminals, coastally moved through tankers and also transferred through cross-country pipeline from Visakhapatnam to Secunderabad via Vijayawada (VVSPL).

VISAKH REFINERY MODERNIZATION PROJECT (VRMP):

VRMP is a brownfield expansion of the Visakh refinery located in Visakhapatnam, India. The Project is being executed by the state-owned Hindustan Petroleum Corporation Limited (HPCL). VRMP will expand the refining capacity of the Visakh refinery from 8.3 million tonne per year to 15 million tonne per year, while enabling the production and distribution of Bharat Stage-VI (BS-VI) – compliant fuels.

BS VI is a new emission standard that will align India’s motor vehicle regulations with European Union (EU) regulations for light-duty passenger cars and commercial vehicles, heavy-duty trucks and buses, and two-wheeled vehicles. The standards is due to be implemented in India from 2020. The VRMP Project is estimated to involve an investment of approximately $3.1bn and is scheduled for completion in 2020.

The objective of the project is to modernize and enhance the capacity of the Visakh Refinery from the present capacity of 8.3 MMTPA to 15.0 MMTPA.

The Project comprise of a new crude unit, Residue up-gradation facilities, revamp of MS and DHT block to produce BS-VI grade MS and HSD , augmentation of Utility systems, integrated effluent treatment system, Offsite facilities, Captive Power Plant along with grid connectivity and other associated facilities. The cost of the project is estimated at Rs.26264 crores.

Details and Technology Upgrades of Visakh Refinery:

HPCL received approval from the Ministry of Environment, Forest, and Climate Change (MoEFCC) for the VRMP in February 2016. The expansion project is being carried out on an area adjacent to the east portion of the existing site. The scope of the project consists of the installation of primary processing units such as a 9 MMTPA Crude Distillation Unit(CDU) replacing one of the three existing CDU's, a 3.3 MMTPA Vacuum Gas Oil Hydrocracker and a 290,000 TPA Naphtha Isomerization Unit (NIU).

An additional 3.1 million tonne per year Solvent De-asphalting (SDA) unit, a 2.5 million tonne per year Slurry Hydrocracker unit (SHCU) and a 96t/day Propylene Recovery unit (PRU) are also outlined in the plans. Other auxiliary units being built under the project include two 113 million tonne per year Hydrogen Generation Units (HGUs), two 360t/day Sulphur Recovery Units (SRUs) including Tail Gas Treatment Units (TGTUs), and a 36,000 tonne per year fuel gas Pressure-Swing Adsorption (PSA) unit. A 300t/h non-hydro-processing Sour-Water Stripper (SWS-I), a 185t/h hydro-processing Sour-Water Stripper (SWS-II), two 540t/h Amine Regeneration Units (ARUs), a 112,000 tonne per year sulphur recovery LPG treating unit will also be installed.

The current Effluent Treatment Plants(ETPs) are set to be replaced by a 1,000 m^3/h Integrated Effluent Treatment Plant (IETP).Various existing units, including a naphtha hydro-treating unit, a continues catalytic reformer and a diesel hydro-treating unit are also being revamped as part of the modernisation. Additionally, a duel fuel-powdered captive power plant (CPP) will be built. The project will also involve the construction of a new utility system comprising a re-circulating sea cooling water system consisting of cooling tower cells and water pumps, as well as a demineralised water system featuring a new reverse osmosis (RO) system and water tanks.

VRMP will implement Honeywell UOP Penex process technology to produce a high-octane gasoline component called isomerate, and a Unicracking hydrocracking unit to produce cleaner burning diesel fuel. The Penex employs high-activity isomerisation and benzene impregnation catalysts from Honeywell UOP. Unicracking processes use highly effective catalysts to yield higher quantities of cleaner-burning fuels from a range of feedstock. The innovative configuration of the fractionation unit ensures effective separation of the products, while minimising capital costs and operating expenses.

The two processes will significantly increase the capability of the refinery to deliver petrol and diesel in compliance with the new BS-VI clean fuels standards. “VRMP will expand the refining capacity of the Visakh Refinery from 8.3 million tonne per year to 15 million tonne per year, while enabling the production of Bharat Stage-VI (BS-VI)-compliant fuels.” The Planned integration of innovative technologies will also enhance gross margins of the refinery by generating high-value transportation fuels from low-value raw materials.

CHAPTER 3

INTRODUCTION TO FLUID CATALYTIC CRACKING (FCC)

Fluid catalytic cracking is a catalytic conversion process for converting heavy gas oils, namely, vacuum distillates into more valuable products such as LPG, gasoline, cycle oils, olefin rich light hydrocarbons that may be further processed to even more valuable products an fuel oils. The decant oil can be used as feedstock for carbon black and needle coke manufacture. The fluid catalytic cracking is a low pressure, intermediate to high temperature process. This process may be designed and operated to achieve either of the processing objectives:

(a) Maximization of middle distillates;
(b) Maximization of LPG and Gasoline.

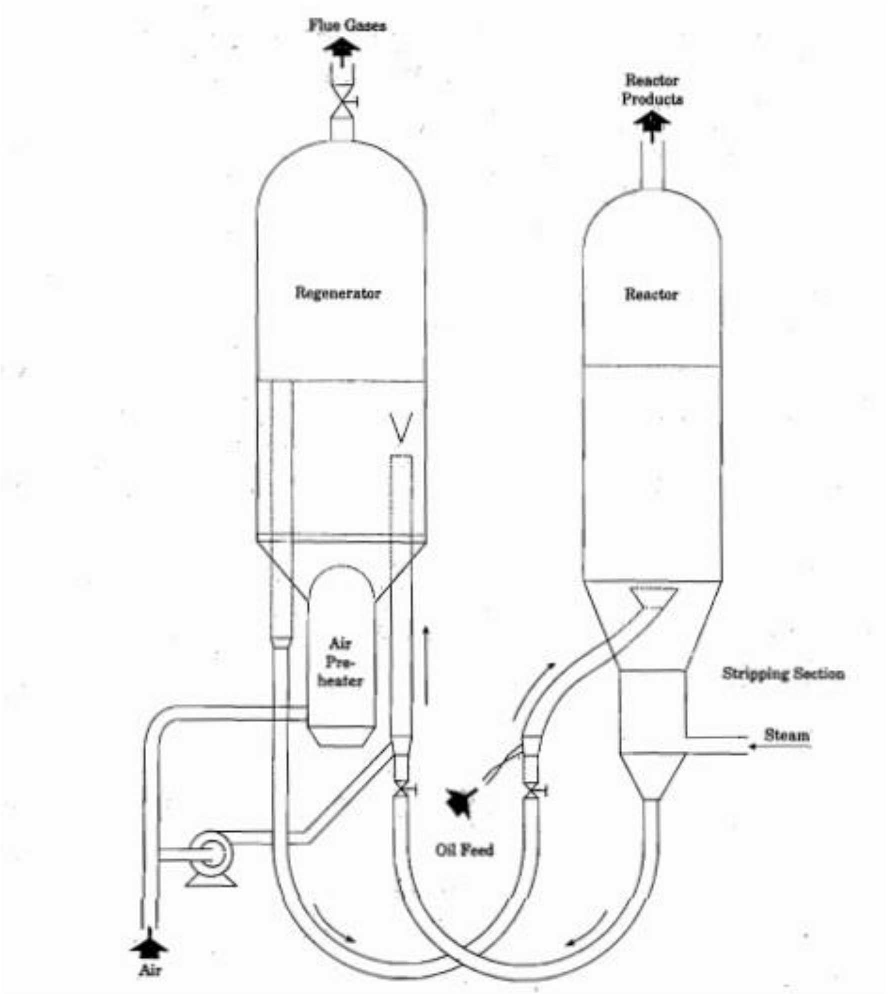
In the former case the catalytic cracker essentially functions as a visbreaker. The transition from one operation to another can be made rapidly without expensive revamps and turn-arounds. This process can accept a wide variety of feedstocks thereby enhancing the total refinery flexibility.

Development of Fluid Catalytic Cracking (FCC):

The fluid catalytic cracking (FCC) process is an outgrowth of a fixed-bed catalytic cracking process developed in France by Eugene Houdry in 1922. The first unit went on stream in the 1930's employing a 10 minute make-regenerate cycle, and using expensive tubular reactors in which molten heat transfer salt was circulated in the shell. The reactors were constructed of expensive alloys to withstand alternate reducing and oxidizing conditions. This type of process, though proved in operation, had the disadvantage of intermittent process and a high initial investment and operating costs.

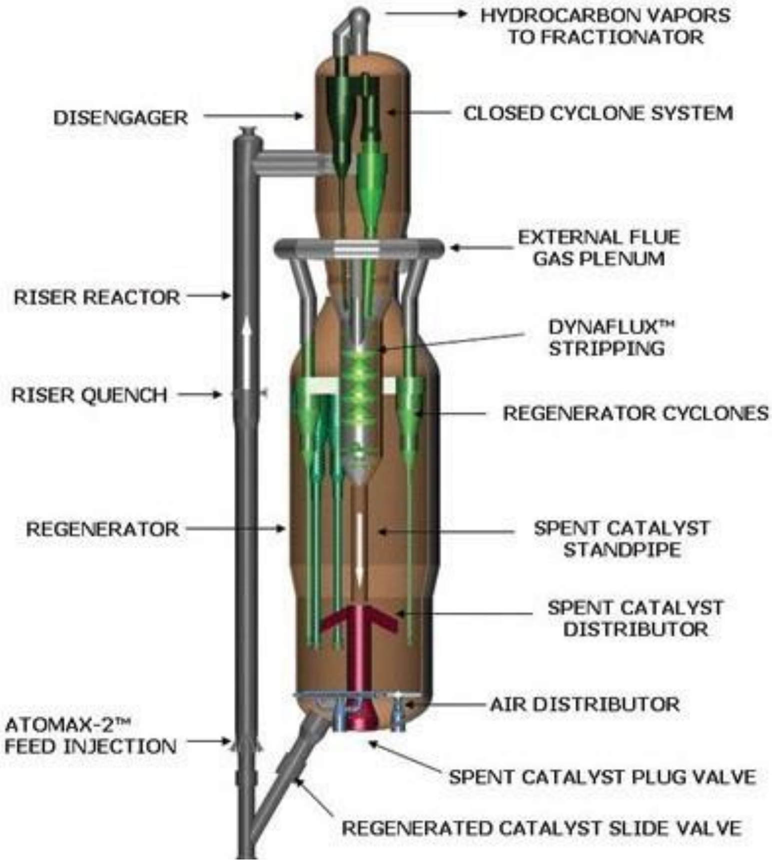
The advantages of a continuous process led to the development of the idea of a moving-bed of catalyst. Examples of this type are Thermo for Catalytic Cracking, the pelletized catalyst was conveyed between the reactor and regenerator by means of bucket elevators. A radical development was made by Standard Oil Co. (New Jersey), M.W.Kellog and U.O.P. in early 1940's in which the catalyst in the form of fine powder was held in suspension in gas streams. It was found that by carefully controlling the catalyst particle size and the velocity of the gas moving through it, a fluidized-bed of catalyst would form which has the properties of a static quantity of liquid. In the fluidized system, finely powdered catalyst is lifted into the reactor by incoming oil which immediately vaporizes upon contact with the hot catalyst and after reaction is complete it is lifted into the regeneration zone. Catalytic crackers using powdered catalyst in this way are known as fluid catalytic cracking units.

The fluid catalytic cracking process has advanced rapidly with the improvements made in each of the areas of catalysts, reactor-regenerator systems and energy recovery. The modern design of catalytic cracking units are described in the figure.



The principal unique feature in this design is the catalyst transfer system. The reactor and regenerator are connected by means of U tubes – one leg of which serves as standpipe and the other as a riser. In order to provide a seal to prevent gases from flowing back from flowing back from one vessel to the other during pressure surges, the air or feedback is not introduced at the bottom of the riser but at a substantial elevation, so that the lower section always contains catalyst in the dense phase. This lower section forms a seal against reverse flow which in other units is provided by the pressure drop over the side valves. The slide valves in Model IV units do not serve to regulate the catalyst circulation rate, as in other units, but they function as safety devices and shut off the lines automatically if there is an imminent danger of reverse flow. The rate of air flow to the spent catalyst transfer line establishes the catalyst circulation rate in the system.

Kellog Orthoflow design is a single vessel type unit. The vessel is divided into two sections, the upper section serving as reactor and stripper. This is known as 'Orthoflow' design. This is shown below.



The catalyst lines are located inside the regenerator and this necessitated the development of special plug valves for controlling catalyst flow. The plug valves consist of conical plugs which can be moved into or out of conical seats, thus decreasing or increasing the cross sectional area through which the catalyst can flow. The feed is injected into the reactor riser through the hollow stem of the regenerator catalyst plug valve.

Cracking in dilute phase gives a better gasoline yield (at constant conversion) than cracking in dense phase. The effective feature of riser crackers-Kellog Orthoflow model F, in addition to dilute phase cracking, is its approach to concurrent flow of catalyst and reacting vapors; there is no backmixing of catalyst and each increment of cracking takes place over catalyst having the lowest carbon deposit and highest effective activity for a given catalyst to oil ratio. Feed enters the riser through nozzles located at the periphery of the riser. Even flow of hydrocarbon on catalyst is realized initially by

(a) ensuring constant flow of catalyst to feed inlet; and
(b) providing a number of feed introduction points.

Maintaining even distribution is achieved by a straight vertical riser with minimum changes in riser diameter. If a diameter change is necessary, a gradual slope is used. Reaction time is controlled by using cyclones at riser exit to separate hydrocarbon vapors from catalyst and discontinue catalytic cracking

Long run times are achieved, on riser crackers, by

(a) smooth and even flow, particularly of catalyst;

(b) limiting velocities below those that cause excessive erosion; and

(c) readily available riser location for easy on-stream maintenance.

Kellog Orthoflow Model F has a two stage regenerator which permits obtaining extremely low carbon on regenerated catalyst levels at low inventories, and, at the same time, promotes high activity maintenance. The first stage is operated at high velocities and low temperatures to promote rapid burning without excessive high particle temperatures. The second stage accomplishes the final burning to very low levels of carbon on catalyst at higher temperatures and lower velocity

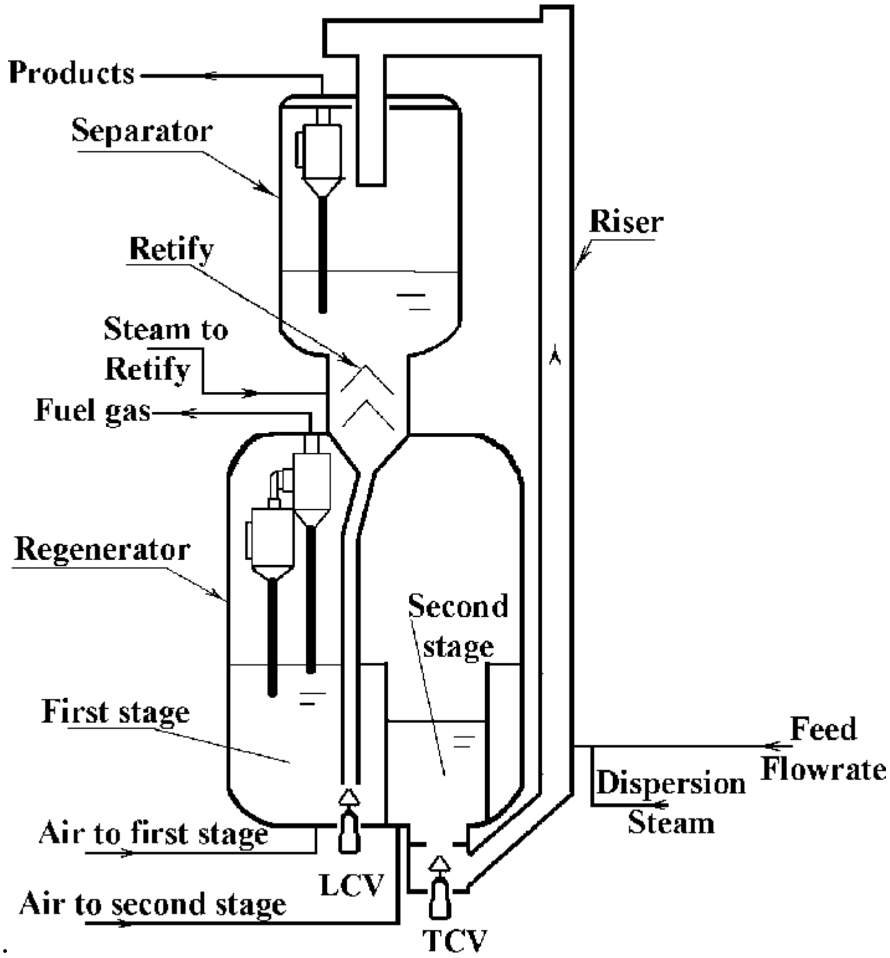


Fig: Kellog Orthoflow Model F-Converter with two-stage regeneration

Technologies Aspects of FCC:

FEEDSTOCK. Vacuum gas oils in the boiling range of 350⁰-550⁰C are used as feedstock in the FCC Process. These are obtained by distillation of atmosphere residue under vacuum. The cut point of vacuum gas oil is controlled to limit the concentration of sulphur, nitrogen and metals in the FCC feed which deactivate catalyst to reduce yield and cause pollution problems due to SO_x and NO_x emissions. Asphaltenes in the feedstock increase the coke yield and may result in high regenerator temperatures.

Other feedstocks may be include atmospheric gas oils boiling below 350⁰C, heavy gas oils obtained from cokers and visbreakers, propane deasphalted oils, etc. Reduced crude may also be included in the FCC feed if its containment levels are not too high.

As a first step to improve distillate yields and eliminate pollution problems, the FCC feed can be upgraded by hydrotreating. Hydrotreating of the FCC feed reduces sulphur, nitrogen and metals contents and thus eliminates pollution problems. The removal of nitrogen prevents deactivation of acidic sites in catalyst by basic nitrogen compounds. The reduction of metals (vanadium and nickel) minimizes dehydrogenation reaction and formation of coke. With appropriate reduction in the levels of containments, the FCC operation can be carried out at higher severities to give better yields of products.

REACTIONS. Catalytic cracking reactions are believed to proceed through the formation of carbonium ions. In the presence of acidic catalyst, carbonium ions are formed either by the removal of a hybride ion from a saturated hydrocarbon or by the addition of a proton to an olefin or aromatic nucleous. Carbonium ions are highly reactive and are responsible for almost all of the reactions occurring in the catalytic cracking process. The formation of carbonium ions depends on the relative stability which decreases in the following order:

Tertiary>Secondary>Primary>Ethyl>Methyl

These carbonium ions undergo a number of reactions which include isomerization, carbon-carbon bond scission, hydrogen transfer and termination.

The reactivity of various types of hydrocarbons having the same carbon number increases in the following order:

Unsubstituted Aromatics < Paraffins < Polymethylaromatics < Naphthenes < Alkylbenzenes (with side chains greater than C_3) < Olefins

CATALYST. The Catalyst used in the catalytic cracking process is a fine powder made up primarily of alumina and silica. The particles of this powder are in the size range of 10-140 microns. In appearance, it might be somewhat analogous to commercial talcum powder. Two types of catalyst are currently significant-amorphous and zeolite.

The synthetic amorphous catalyst has been in use since 1942 though over the years it was substantially improved. It was developed as an outgrowth of the earlier natural catalyst which was prepared by acid treating of certain alloys. The synthetic amorphous catalyst is formed by the precipitation of silica to form a highly porous structure. On this porous structure, alumina is added as a surface coating. Active sites which are acidic in nature are formed at the silica-alumina interface and reaction takes place at these sites with the available hydrogen.

Zeolites catalysts came into use in FCC during the early to mid-1960's. They are formed through the reaction of reactive forms of alumina and silica. The Reaction conditions are controlled so that the product of the reaction will be a caged like crystalline structure.

CHAPTER-4

PROCESS PRINCIPLE & PROCESS CHEMISTRY

4.1. FCC Process Principle:

The Fluid Catalytic Cracking (FCC) process converts heavy crude oil fraction into lighter, more valuable hydrocarbon products at high temperature and moderate pressure in the presence of a finely divides silica / alumina based catalyst. In the course of cracking large hydrocarbon molecules into smaller molecules, a nonvolatile carbonaceous material, commonly referred to as coke, is deposited on the catalyst. The coke laid down on the catalyst acts to deactivate the catalytic cracking activity of the catalyst by blocking access to the active catalytic sites. In order to regenerate the catalytic activity of the catalyst, the coke deposited on the catalyst is burned off with air in the regenerator vessel.

One of the important advantages of fluid catalytic cracking is the ability of the catalyst to flow easily between the reactor and the regenerator when fluidized with an appropriate vapor phase. In FCC units the vapor phase, on the reactor side is vaporized hydrocarbon and steam, while on the regenerator side the fluidization media is air and combustion gases. In this way, fluidization permits hot regenerator catalyst to contact fresh feed; the hot catalyst vaporizes the liquid feed and catalytically cracks the vaporized feed to form lighter hydrocarbon products. After the gaseous hydrocarbons are separated from the spent catalyst, the hydrocarbon vapor is cooled and then fractionated into the desired product streams. The separated spent catalyst flows via steams fluidization from the reactor to the regenerator vessel where the coke is burned off the catalyst to restore its activity. In the course of burning the coke a large amount of heat is liberated. Most of this heat of combustion is absorbed by the regenerated catalyst and is carried back to the reactor by the fluidized regenerated catalyst to supply the heat required to drive the reaction side of the process. The ability to continuously circulate fluidized catalyst between the reactor and the regenerated allows the FCC to operate efficiently as a continuous process.

A modern UOP designed FCCU consists of three sections Catalyst section, Fractionation Section & Gas Concentration Unit which operates together in an integrated manner.

A. Catalyst Section

The Catalyst Section consists of the Reactor and Regenerator which together with the Stand pipes and Riser from the “Catalyst Circulation Circuit”.

The fresh feed and recycle streams known as combined feed enter the unit at the base of the riser where they are vaporized and raised to the reactor temperature by the hot catalyst. The mixture of oil vapor and catalyst travels up the riser into the reactor. The gas oil starts to crack immediately when it contacts catalyst in the riser and continues until the oil vapors are disengaged from the catalyst in the reactor. The cracked products in the vapor phase go to fractionator through the reactor vapor line. Coke is deposited on the circulating catalyst in the reaction zone. The spent Catalyst flows from the reactor to the regenerator where coke is burnt off. The heat of combustion raises the Catalyst temperature to 670-721 °C range and supplies heat, most of which is transferred to the change in the riser. The coke deposited on the circulating catalyst in the reactor is burnt off in the regenerator using air from Air Blower. Catalyst gets separated from the hydrocarbon vapors in the reactor and from flue gases in the regenerator by means of set of cyclones provided at the top section of reactor and regenerator.

The flue gas coming out from regenerator is a mixture of CO and CO_2 gases which are burnt in CO Boiler for producing steam. The flue gases can be vented to atmosphere through stack bypassing CO Boiler.

B. Fractionation Section

In the fractionation section, the reactor vapors are fractionated into the products clarified oil, Heavy Cycle oil, Light Cycle Oil, Unstabilized Gasoline and Wet Gas. The unit is also designed to produce High pressure and Medium Pressure steam in the exchangers by transferring waste heat from circulating streams, of fractionation section. Unstabilized Gasoline and Wet Gas after compression are sent to Gas Concentration Unit for further separation.

C. Gas Concentration Section

In Gas Concentration section Gasoline (CRN), LPG and Sour Fuel Gas (ATG) are separated. The yield and composition of each stream is optimized for maximum product value. Gasoline and LPG are sent to FCC-NHT and NLU units respectively for further treatment. Sour fuel gas from GCU joins Refinery fuel gas system after the removal of $H_2S$and mercaptans in FG-AAU (SRU). Gasoline streams produced in FCCUsis generally of high octane number.

5.2. FCC PROCESS CHEMISTRY

FCC brings together heavy hydrocarbon feed with fluidized circulating catalyst. During cracking reactions, carbon and hydrogen (i.e., Coke) is deposited on catalyst which reduces the catalyst activity. This coke is burnt in the regenerator, which generates the heat to sustain and continue the reaction and the catalyst activity is also restored.

Gas oil is portion of crude oil that boils in 330-550 ^{o}C range and contains a mixture of Paraffins, Naphthenes and Aromatics. About 50% of the feed consists of saturated molecules (Paraffins and Naphthenes) and remaining molecules contain at least one aromatic ring. Aromatic molecules have saturated side chains and/or naphthene rings attached to them.Paraffins, Naphthenes and sidechains can crack to smaller molecules whereas aromatic rings cannot be cracked.

Type of molecules in cracked products
Naphtha ------------------------ Paraffins, Naphthenes, Olefins & 1-Ring aromatics
Heating Oil -------------------- 2- and 3- Ring aromatics
Bottoms and Coke ----------- 4+ Ring Aromatics

List of Main FCCU Reactions

Cracking reactions

- Paraffins - Olefins Olefins+ Paraffins
- Naphthenes - Olefins
- Aromatics - Un-substituted aromatics+ Olefins
- Olefins - Smaller LPG Olefins

Hydrogen Transfer

- Olefins - Paraffins
- Branched olefins - Branched Paraffins
- Olefins &Naphthenic - Paraffins + Aromatics

Dehydrogenation

- Naphthenes - Cyclo Olefins & aromatics
- Aromatics - Heavy Heterocyclic Aromatics (coke)
- Olefins - Coke (Net Reduction)

Isomerization

- Olefins - Branched Olefins

Cyclization

- Olefins - Naphthenes

Alkylation

- Aromatics - Aromatics Heavy Heterocyclic Aromatics (coke)
- Aromatics - Different alkyl aromatics

Mentioned above is a list of major reactions that take place in reactor apart from numerous other side reactions which provides innumerable ways of combining FCC with downstream units to produce valuable products. This is perhaps the charm of FCCU which encourages the researchers around the globe.

CHAPTER-5

OPERATION AND MONITORING OF FCC

- The proper way to monitor the performance of a Fluid Catalytic Cracker is by periodic material and heat balance surveys on the unit. By carrying out these tests frequently, one can collect, trend, and evaluate the unit operating data. Additionally, meaningful technical service to optimize the unit operation should be based on regular test runs.
- Understanding the operation of a catalytic cracker also requires in-depth knowledge of the unit's heat balance. Any changes to feedstock quality, operating conditions, catalyst, or mechanical configuration will impact the heat balance. Heat balance is an important tool in predicting and evaluating the changes that will affect the quantity and the quality of FCC products. Finally, before the unit can produce a single barrel of product, it must circulate catalyst smoothly and therefore one must be quite familiar with the dynamics of pressure balance.

➢ Material balance

➢ Heat balance

➢ Pressure balance

5.1 Material Balance:

Complete data collection should be carried out weekly. Since changes in the unit are continuous,regular surveys permit distinction among the effects of feedstock, catalyst, and operating conditions.An accurate assessment of a cat cracker operation requires reliable plant data. A reasonable weight balance should have a 98%e102% closure.In any weight balance exercise, the first step is to identify the input and output streams.

One of the key objectives of conducting the mass balance exercise is to determine the composition of products leaving the reactor. The reactor effluent vapors entering the main fractionator containhydrocarbons, steam, and inert gases. By weight, the hydrocarbons in the reactor overhead stream are equal to the fresh feed plus any recycle minus the portion of the feed that was converted to coke. The main sources of steam in the reactor vapors are: lift steam to the riser, atomization steam to the feed nozzles, reactor dome steam, and stripping steam. Some FCC units may purposely inject water into the feed injection system as part of heat removal from the regenerator. Depending on the reactor pressure and catalyst circulation rate, approximately 25% e 50% of the stripping steam is entrained with the spent catalyst flowing to the regenerator and should be deducted.

5.2 Testing methods:

The material balance around the riser requires the reactor effluent composition. Two techniques are used to get this composition. Both techniques require that the coke yield be calculated.The first technique is to draw an envelope with the reactor effluent as the inlet stream and the product flows as the outlet streams. Included in this envelope must be any external streams that are entering into the main fractionator and/or FCC gas plant circuits. The reactor yields and its composition are determined by subtracting the products from the main fractionator and gas plant, from the external streams. This is the method practiced by most refiners.

5.3 Data collection:

1. The duration of a test run is usually 12-24 h.
2. Operating parameters should be specified. It should be documented which constraints (i.e. blower, wet gas compressor, etc.) the unit is operating against.
3. The sample taps must be bled adequately before samples are collected. A reliable flue gas analyzer that display not only O_2, but also CO and CO_2 .; an extra sample can be collected. The laboratory should retain the unused samples until all analyses are verified.
4. Sponge or secondary absorber off-gas and C_3/C_4 samples must be collected upstream of the amine treaters (if possible) to ensure proper fractions of H_2S is reported.
5. Pertinent operating data must be collected.

5.4 Mass balance calculations:

1. The orifice plate meter factors should be adjusted for actual operating parameters. For liquid streams, the flow meters should be adjusted for # API gravity, temperature, and viscosity. For gas streams, the flow rate should be adjusted for the operating temperature, pressure, and molecular weight.

2. Chromatographs of each stream must be normalized to 100%. The GC of the off-gas m3 must include accurate analysis of hydrogen sulfide (H_2S)

3. The coke yield should be calculated using air rate and flue gas composition.
4. The flow rate of each stream should be converted to weight units.
5. The quantity of inert gases and extraneous streams should be subtracted from the FCC gas plant products.
6. The raw mass balance should be reported, including the error. Then the feed/products should be normalized to 100%. The error will be distributed in proportion to flow rates or a known inaccurate meter will be adjusted.
7. Gasoline, LCO and slurry flow rates will be adjusted to standard cut points.
8. The feed characterization correlations discussed in Chapter 3 should be used to determine thecomposition of fresh feed.

5.5 Analysis of results:

1. The yields and quality of the desired products should be reported and compared with the unit targets.

2. The GC analyses must make sense

3. The results of this test run should be compared with the results of previous test runs; any significant changes in the yields and/or operating parameters should be highlighted.

4. The final step is to perform simple economics of the unit operation and make recommendations that improve unit operation short and long term.

5.6 The mass balance is performed as follows:

1.Identification of the input and output streams used in the overall mass balance equation.

2.Calculation of the coke yield.

3.Conversion of the flow rates to weight units, e.g., lb/h or kg/h.

4.Normalization of .the data to obtain a 100% weight balance.

5.Determination of the component yields.

6.Adjustment of the gasoline, LCO, and slurry oil yields to standard cut points

5.7 Coke yield calculations:

A portion of the feed is converted/deposited to coke in the riser/reactor housing and catalyst stripper. This coke is carried into the regenerator with the spent catalyst. The combustion of the coke produces H_2O, CO, CO_2, SO_2, and traces of NOx. To determine the coke yield, the amount of dry air to the regenerator and the analysis of the regenerator flue gas are needed. It is essential to have an accurate analysis of the flue gas. The hydrogen content of coke relates to the amount of volatile hydrocarbons that are carried under with the spent catalyst into the regenerator, and is an indication of the reactor-stripper performance. Example 8.1 shows a step-by-step calculation of the coke yield.

5.8 Heat balance:

A catalytic cracker is a coke rejection process. It continually adjusts itself to stay in heat balance. This means that the reactor and regenerator heat flows must be equal Simply stated, the unit produces and burns enough coke to provide energy to:

- Vaporize fresh feed and any recycle streams
- Increase the temperature of the fresh feed, recycle, and all the steam to the riser m from their preheated states to the reactor temperature.
- Provide the endothermic heat of cracking.
- Increase the temperature of the combustion air from the blower discharge temperature to the regenerator dilute phase temperature.

- Make up for heat losses from the reactor and regenerator to the surroundings.
- Provide for miscellaneous heat sinks, such as stripping steam and catalyst cooling.
- A heat balance can be performed around the reactor, around the stripper-regenerator, and as an overall heat balance around the reactor-regenerator. The stripper-regenerator heat balance can be used to calculate the catalyst circulation rate and the catalyst-to-oil ratio.

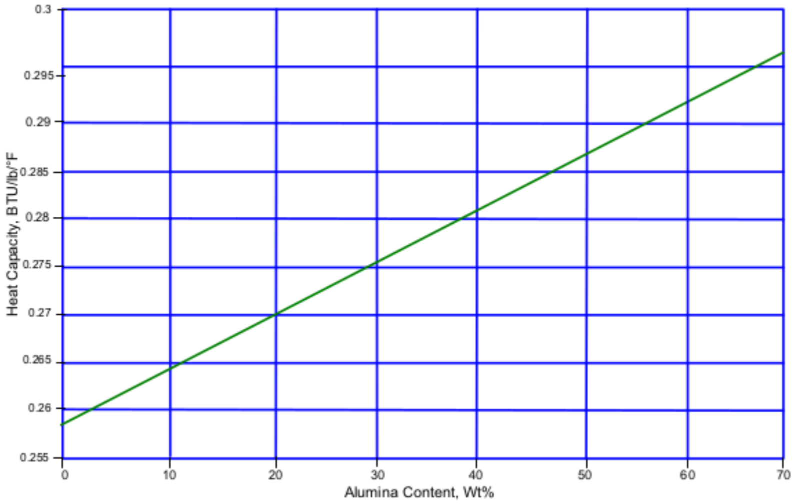
5.8.1 Heat balance around stripper-regenerator:

If a reliable spent catalyst temperature is not available, the stripper is included in the heat balance

The combustion of coke in the regenerator satisfies the following

heat requirements:

- ➔ Heat to raise air rate from the blower discharge temperature to the regenerator dilute phase temperature.
- ➔ Heat to desorb the coke from the spent catalyst.
- ➔ Heat to raise the temperature of the stripping steam to the reactor temperature.
- ➔ Heat to raise the coke on the catalyst from the reactor temperature to the regenerator dense phase temperature.
- ➔ Heat to raise the coke products from the regenerator dense temperature to flue gas temperature.
- ➔ Heat to compensate for regenerator heat losses.
- ➔ Heat to raise the spent catalyst from the reactor temperature to the regenerator dense phase temperature.



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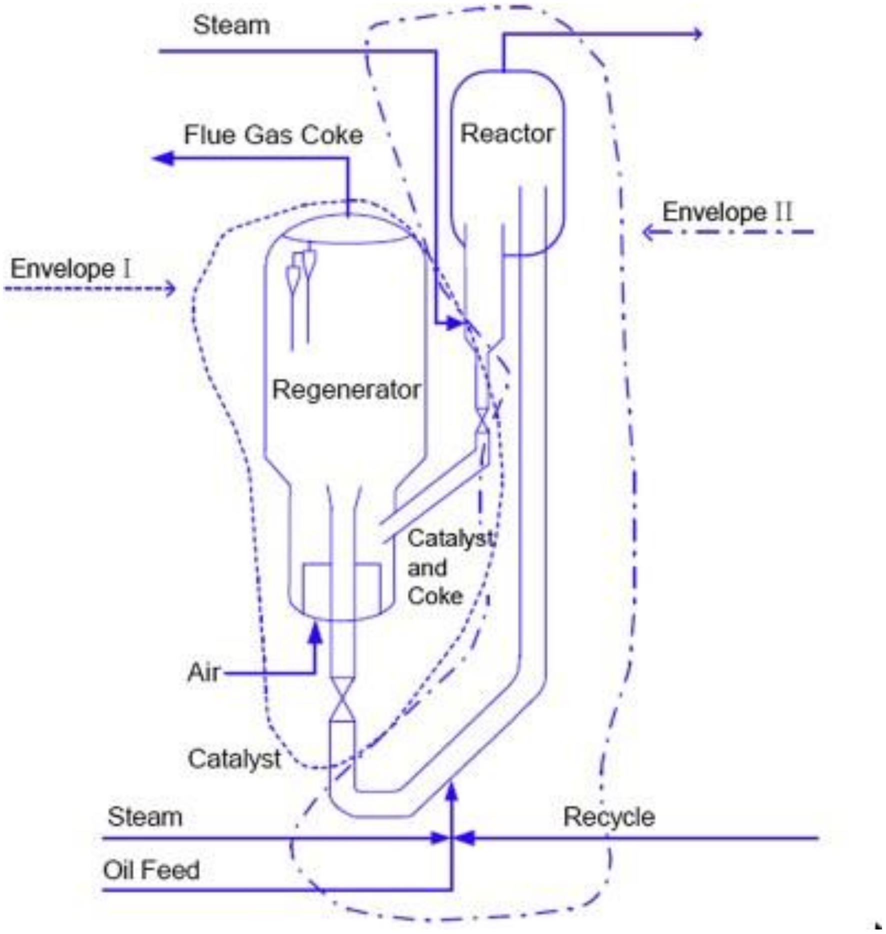
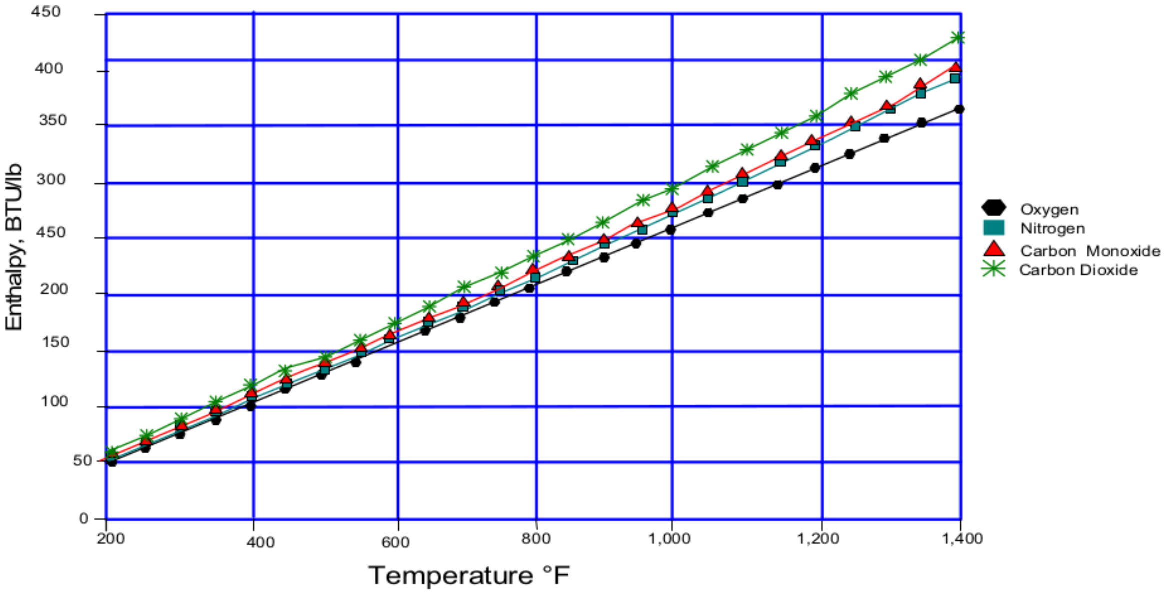
Heat capacity of FCC catalyst as function of alumina content

Reactor and Regenerator:



enthalpies of FCC and flue gas components

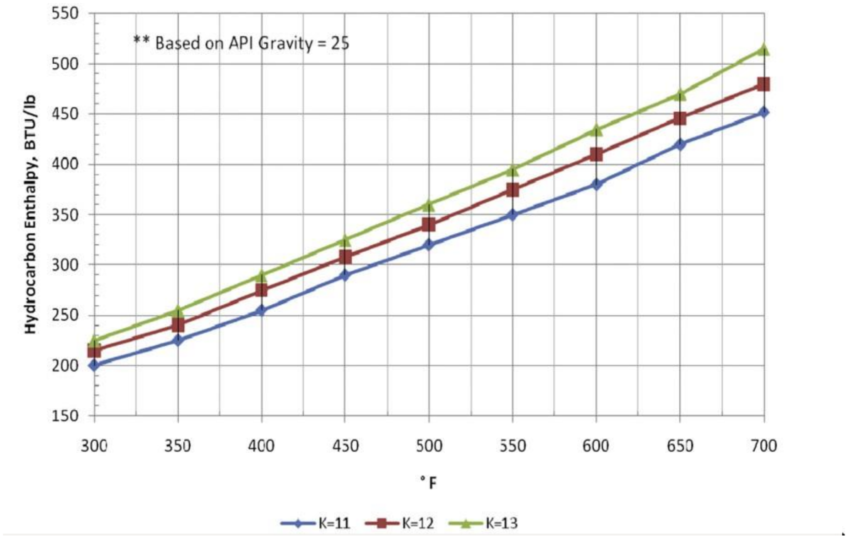


5.8.2 Reactor Heat Balance:

The hot regenerated catalyst supplies the bulk of the heat required to vaporize the liquid feed (and any recycle), to provide the overall endothermic heat of cracking, and to raise the temperature of dispersion steam and inert gases to the reactor temperature.

Heat in	Heat out
Fresh feed	Reactor vapors
Recycle	Flue gas
Air	Losses
Steam	

It is calculated as the net heat from the heat balance divided by the feed flow in weight units. This approach to determining the heat of reaction is acceptable for unit monitoring. However, in designing a new cat cracker, a correlation is needed to calculate the heat of reaction. The heat of reaction is needed to specify other operating parameters, such as preheat temperature. Depending on conversion level, catalyst type, and feed quality, the heat of reaction can vary from 120 Btu/lb to 220 Btu/lb. In the unit, the heat of reaction is a useful tool. It is first an indirect indication of heat balance accuracy. Trending the heat of reaction on a regular basis provides insight into reactions occurring in the riser and the effects of feedstock and catalyst changes.



Hydrocarbon liquid enthalpies at various Watson K factors.

5.9 Analysis of results:

Once the material and heat balances are complete, a report must be written. It will first present the data.It will then discuss factors affecting product quality and any abnormal results. The report needs to discuss the key findings and recommendations to improve unit operation. the feed characterizing correlations are used to determine composition of the feedstock. The results show that the feedstock is predominantly paraffinic, i.e., 61.6% paraffins, 19.9% naphthenes, and 18.5% aromatics. Paraffinic feedstocks normally yield the most gasoline with the least octane. This confirms the relatively high FCC gasoline yield and low octane observed in the test run. This is the kind of information that should be included in the report. Of course, the effects of other factors such as catalyst and operating parameters will also affect the yield structure.The coke calculation showed the hydrogen content to be 9.9 wt%. Every effort should be made to minimize the hydrogen content of the coke entering the regenerator. The hydrogen content of a well-stripped catalyst is the range of 5 wt% to 6 wt%. A 9.9 wt% hydrogen in coke indicates either poor stripper operation and/or erroneous flue gas analysis.

5.10 Pressure balance:

Pressure balance deals with the hydraulics of catalyst circulation in the reactor/regenerator circuit. The pressure balance starts with conducting a single-gauge pressure survey of the reactor-regenerator circuits. The overall objective is

➔ To ensure steady catalyst circulation is achieved

➔ To maximize catalyst circulation

➔ To maximize the available pressure drop at the slide valves; and

➔ To minimize the loads on the blower and the wet gas compressor.

To A clear understanding of the pressure balance is extremely important in “squeezing” the most out of a unit. Incremental capacity can come from increased catalyst circulation or from altering the differential pressure between the reactor-regenerator to “free up” the wet gas compressor or air blower loads. One must know how to manipulate the pressure balance to identify the “true” constraints of the unit. Using the drawing(s) of the reactor-regenerator, the unit engineer must be able to go through the pressure balance and determine whether it makes sense. He or she needs to calculate and estimate pressures, densities, pressure buildup in the standpipes, etc. The potential for improvements can be substantial.

5.10.1 Basic fluidization principals:

A fluidized catalyst behaves like a liquid. Catalyst flow occurs in the direction of a lower pressure. The difference in pressure between any two points in a bed is equal to the static head of the bed between these points; multiplied by the fluidized catalyst density, but only if the catalyst is fluidized. FCC catalyst can be made to flow like a liquid but only if the pressure force is transmitted through the catalyst particles and not the vessel wall. The catalyst must remain in a fluidized state as it makes a loop through the circuit. To illustrate the application of the above principals, the role of each major component of the circuit is discussed in the following sections, followed by an actual case study.

5.10.2 Major components of the reactor-regenerator circuit :

The major components of the reactor-regenerator circuit that either produce or consume pressure are as follows:

- Regenerator catalyst hopper
- Regenerated catalyst standpipe
- Regenerated catalyst slide (or plug) valve
- Riser
- Reactor-stripper
- Spent catalyst standpipe
- Spent catalyst slide (or plug) valve.

5.10.2.1 Regenerator catalyst hopper :

In some FCC units, the regenerated catalyst flows through a hopper prior to entering the standpipe.The hopper is usually internal to the regenerator. The hopper is intended to provide sufficient residence time for the regenerated catalyst to be deaerated before entering the standpipe. This causes the catalyst entering the standpipe to have its maximum flowing density, the higher the catalyst flowing density, the greater the pressure buildup in the standpipe. In some FCC designs, theregenerated catalyst hopper is external with fluffing aeration to control the catalyst density entering the standpipe.

5.10.2.2 Regenerated catalyst standpipe:

The standpipe's height provides the driving force for transferring the catalyst from the regenerator tothe reactor. The elevation difference between the standpipe entrance and the slide valve is the source of this pressure buildup. For example, if the height difference is 30 feet (9.2 m) and the catalyst flowing density is 40 lb/ft 3 (641 kg/m 3), the pressure buildup is:

$$\text{Pressure gain} = 30 \text{ ft} \times \frac{40 \text{ lb}}{\text{ft}^3} \times \frac{1 \text{ ft}^2}{144 \text{ in}^2} = 8.3 \text{ psi}(57 \text{ kPa})$$

The key to obtaining maximum pressure gain is to keep the catalyst fluidized over the entire length of the standpipe. Longer standpipes will require external aeration. This aeration compensates for compression of the entrained gas as it travels down the standpipe. Aeration should be added evenly along the length of the standpipe. In shorter standpipes sufficient flue gas is often carried down with the regenerated catalyst to keep it fluidized and supplemental aeration is unnecessary. Over-aeration leads to unstable catalyst flow and must be avoided. Aside from proper aeration, the flowing catalyst must contain sufficient 0e40 mm fines, as well as minimum amount of 150 mm particles to avoid de-fluidization.

5.10.2.3 Regenerated catalyst slide valve:

The purpose of the regenerated catalyst slide valve is threefold: to regulate the flow of the regenerated catalyst to the riser, to maintain pressure head in the standpipe, and to protect the regenerator from a flow reversal. Associated with this control and protection is usually a 1 psi to 8 psi (7 kPae55 kPa) pressure drop across the valve.

5.10.2.4 Riser:

The hot-regenerated catalyst is transported up the riser and into the reactor-stripper. The driving force to carry this mixture of catalyst and vapors comes from a higher pressure at the base of the riser and the low density of the catalyst/vapor mix. The large density difference between the fluidized catalyst on the regenerator side (approximately 40 lb/ft 3) and the mixture of cracked hydrocarbon vapors and catalyst on the riser side (approximately 1 lb/ft 3) is what creates the catalyst circulation from the regenerated catalyst slide valve into the reactor housing. As for the pressure balance, this transported catalyst results in a pressure drop in a range of 5 psi to 9 psi (35 kPae62 kPa). This pressure drop is due to the static head of the catalyst from downstream of the slide valve to the feed nozzles, the static head of the catalyst in the riser, friction and acceleration losses from the catalyst/vapors within the riser and its termination device. In an existing riser, operating changes, such as higher catalyst circulation or lower vapor velocity, can affect the density of reaction mixture and increase the pressure drop. Thiswill affect the slide valve differential pressure and operating percent opening.

5.10.2.5 Reactor-stripper:

The catalyst bed in the reactor-stripper is important for three reasons:

- To provide enough residence time for proper stripping of the entrained hydrocarbon vapors
- prior to entering the regenerator.
- To provide adequate static head for flow of the spent catalyst to the regenerator.
- To provide sufficient backpressure to prevent reversal of hot flue gas into the reactor system.

Assuming a stripper with a 20 ft bed level and a catalyst density of 40 lb/ft 3 , the static pressure is:

$$20\ \text{ft} \times \frac{40\ \text{lb/ft}^3}{144\ \text{in}^2/\text{ft}^2} = 5.5\ \ \text{psi}$$

$$\frac{6\ \ \text{m} \times 640\ \ \text{kg/m}^3}{10,197\ \ \text{kg/m}^2/\text{bar}} = 3.8\ \ \text{bar}$$

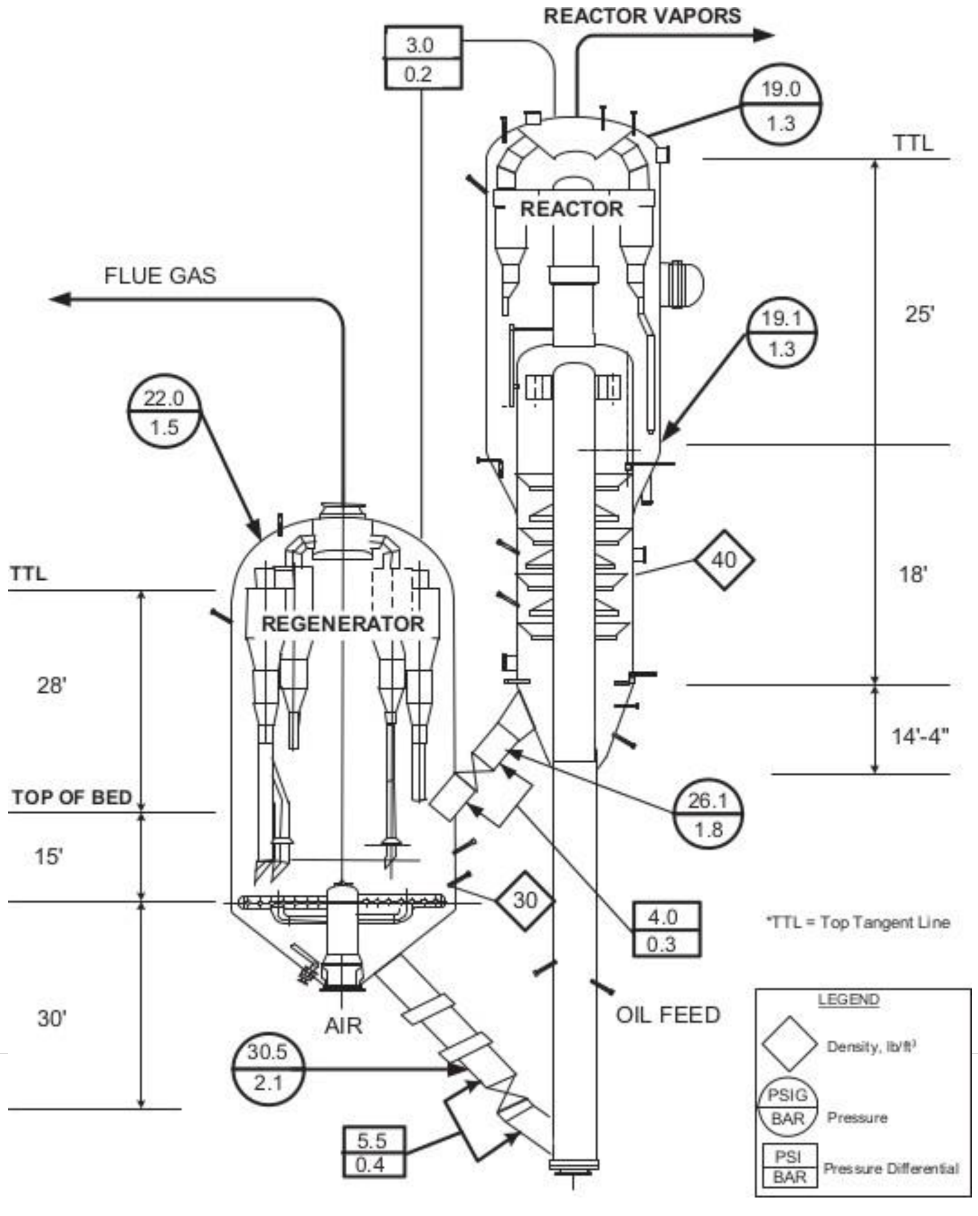
5.10.2.6 Spent catalyst standpipe:

From the bottom of the stripper, the spent catalyst flows into the spent catalyst standpipe. Sometimes the catalyst is partially defluidized in the stripper cone. To counter this, “dry” steam is usually added (through a distributor) to fluidize the catalyst prior to entering the standpipe. The loss of fluidization in the stripper cone can cause a buildup of dense phase catalyst along the cone walls. This buildup can restrict catalyst flow into the standpipe, causing erratic flow and reducing pressure buildup in the standpipe. Like the regenerated catalyst standpipe, the spent catalyst standpipe may require supplemental aeration to obtain optimum flow characteristics. “Dry” steam is the usual aeration medium.

5.10.2.7 Spent catalyst slide or plug valve:

The spent catalyst slide valve is located at the base of the standpipe. It controls the stripper bed level and regulates the flow of spent catalyst into the regenerator. As with the regenerated catalyst slide valve, the catalyst level in the stripper generates pressure as long as it is fluidized. In some of the earlier FCC units, spent catalyst is transported into the regenerator using 50%e100% of the total air to the regenerator. The minimum carrier air velocity to the spent catalyst riser is usually in the range of 30 ft/s (9.1 m/s) to prevent catalyst slumping.

Reactor and Regenerator:



Summary:

The only proper way to evaluate the performance of a catalytic cracker is by conducting a material and heat balance. One balance will tell where the unit is; a series of daily or weekly balances will tell where the unit is going. The heat and weight balance can be used to evaluate previous changes or predict the result of future changes. Material and heat balances are the foundation for determining the effects of operating variables.

The material balance test run provides a standard and consistent approach for daily monitoring. Itallows for accurate analysis of yields and trending of unit performance. The reactor effluent can be determined by direct sampling of the reactor overhead line or by conducting a unit test run.The heat balance exercise provides a tool for in-depth analysis of the unit operation.

Heat balance surveys determine catalyst circulation rate, delta coke, and heat of reaction. The procedures described in this chapter can be easily developed and programmed into a spreadsheet to calculate the balances on a routine basis.

The pressure balance provides an insight into the hydraulics of catalyst circulation. Performing pressure balance surveys will help the unit engineer identify “pinch points”. It will also balance two common constraints: the air blower and the wet gas compressor.

CHAPTER-6

Troubleshooting Most Common FCC Problems

It would be impossible to determine what specific problems are most usual to the Fluid Catalytic Cracking Unit.There are more than 20 variations in mechanical designs of the Regenerator and Reactor still operating throught the world and countless technology component designs that have been applied either from licensor development or refiner internal conception and fabrication. However ,if a broad view is applied to the common elements of all FCCU, then it is possible to define and resolve the major problems that can be encountered regardless of unit configuration.

6.1. The Cavitating Pump

The FCC main column pump for heavy naphtha sidedraw was cavitating during the winter.Lack of sufficient Net Positive Suction Head(NPSH) was suspected, and preliminary plans for a new low NPSH pump were in progress.Typically ,hotter temperatures aggravate pump cavitation. The first clue to the true cause was use of live steam in HN stripper; low temprature strippers, eg in atmospheric towers, typically use indirect stripping with pumparound reboilers. Second clue was the fact that the steam was from the low pressure steam header ,barely superheated.And the final steam was turned way down or even off in the winter because of product planning complaints on free water in the naphtha tankage.

Providing a new pump with lower required NPSH would solve the proximate symptom of cavitation, removing the damage to the pump.

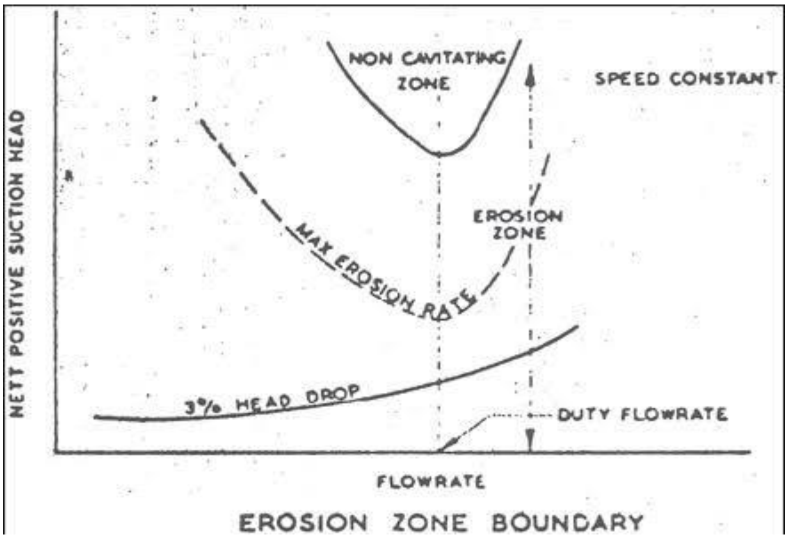


Figure: Safe NPSH margin for a centrifugal pump

6.2. The Breaking Compressor

For most FCC units, the wet gas compressor is centrifugal, but in this smaller refinery, a reciprocating (positive displacement) compressor handled the duty. At least when it was working, and not shut down to replace the piston rings, which was occurring too often since the last turnaround when the FCC unit was expanded in throughput. Liquid droplets were damaging the rings. Buying expensive high-grade metallurgy rings was under consideration. Checking the entry section of the compressor, using an endoscope through the "hand-hole" that was opened during a shutdown for repairs, I could see tracks of liquid condensate marked on the side of the section. Yet the knock-out pot in front of the compressor had also been recently modified, resized and demister packing installed for the feedrate expansion project. In fact, the demister had been resized conservatively, to handle even more than the anticipated maximum wet gas flow rate, and specified as "Impaction" style, to permit use of the low volume of packing allowed by the high velocity permitted.

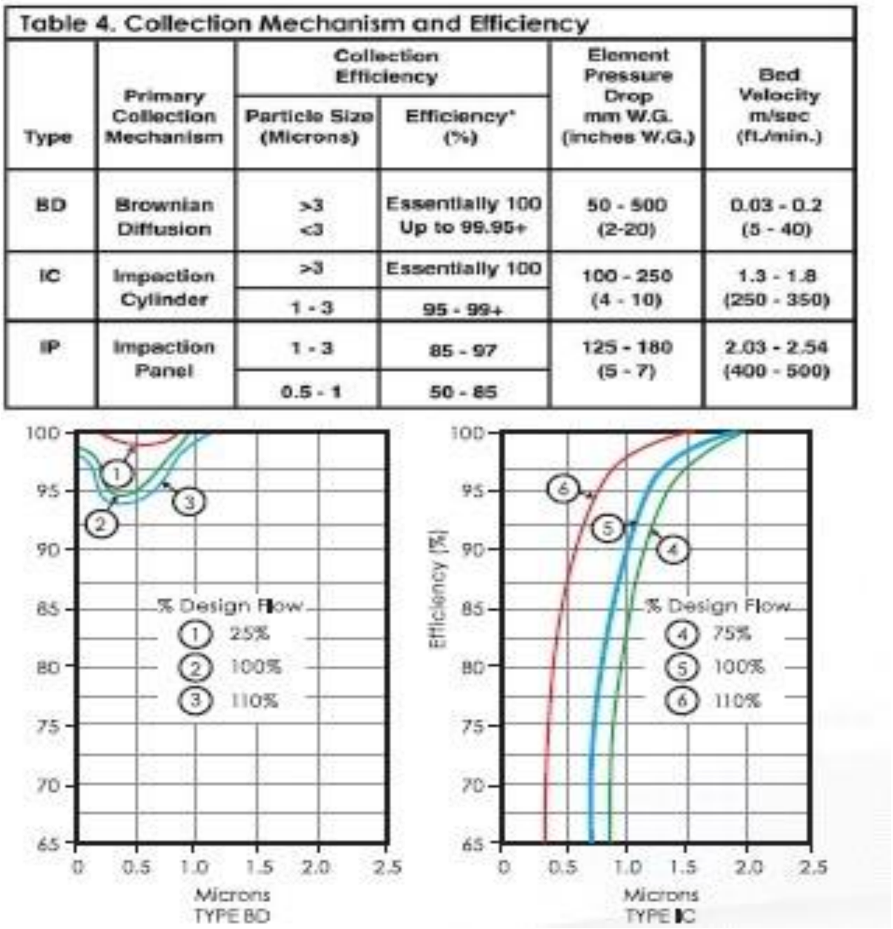


Table 4. Collection Mechanism and Efficiency

Type	Primary Collection Mechanism	Collection Efficiency		Element Pressure Drop mm W.G. (inches W.G.)	Bed Velocity m/sec (ft./min.)
		Particle Size (Microns)	Efficiency* (%)		
BD	Brownian Diffusion	>3 <3	Essentially 100 Up to 99.95+	50 - 500 (2-20)	0.03 - 0.2 (5 - 40)
IC	Impaction Cylinder	>3	Essentially 100	100 - 250 (4 - 10)	1.3 - 1.8 (250 - 350)
		1 - 3	95 - 99+		
IP	Impaction Panel	1 - 3	85 - 97	125 - 180 (5 - 7)	2.03 - 2.54 (400 - 500)
		0.5 - 1	50 - 85		

Table:Collection mechanism and efficiency.

Unfortunately, the "conservative" design meant that the KO pot was operating all the time below the design demister flow rate (and most of the time below the maximum rate for the expansion project). That translates to efficiencies of droplet collection at or below curve 4 above, which was particularly low with the finer droplets. Returning to the original project basis for design and switching to "Brownian Diffusion" style of demister pad, the surface area was greatly increased, but the efficiency was much improved, and almost completely insensitive to turndown flow rates. Higher collection efficiencies than the other style demister were achievable even above project max rates (compare curve 3 with curve 6). A general lesson can also be seen here: Conservative design does not always mean over-design, especially in such items as cyclones (efficiency drops rapidly below design rate) and exchangers in fouling service (low flow rates promote the very fouling feared in the first place).

6.3. The Sodium from nowhere

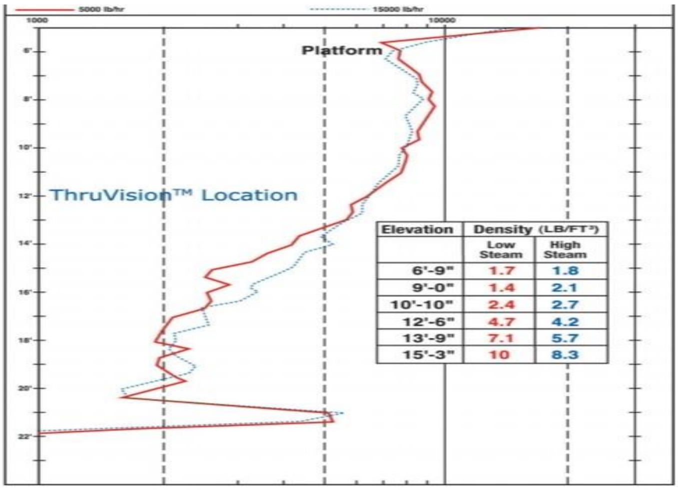
So once again the sodium level was climbing on our FCC e-cat, with predictable effects on catalyst activity and unit performance. We had seen this in prior years, but the symptoms went away on their own, with little to do but maintain an aggressive makeup rate to counteract the impacts. Checking all the feed components (the usual suspect in sodium excursions) showed no evidence of excess sodium, the desalters and vacuum tower overflash were in order, there was no resid processing at the time, fresh catalyst sodium levels were in order, and would not have impacted catalyst activity anyway (high sodium on fresh catalyst hurts octane, but not activity).

The heavy, dense mists were salt-laden, and pulled in to the MAB suction, depositing on the regenerator e-cat for up to several months each year when the weather patterns “allowed”. No clean solution, but at least we could set our catalyst makeup strategy based on the weather-man’s projections.

6.4. Poor Reactor Performance

Poor reaction results occur when conversion and liquid product yields do not match expected or predicted results. This may be due to catalyst contamination and activity but can also be related to other factors in the FCC Riser such as feed and catalyst distribution. Any element that adversely influences delta coke and catalyst circulation rate will bring about a loss of converted product and selectivity. Many FCC operations have excellent catalyst activity maintenance and low contaminants but still experience relatively poor conversion. In these cases, Riser hydrodynamics can be the primary reason. Poor distribution of feed into the catalyst or poor distribution of catalyst into the feed can be crucial to achieving optimum yields.

To achieve optimum yields, it is desirable to have uniform catalyst density in the Riser. If the catalyst is poorly mixed with the feed, areas will be formed in the Riser where there is either too much feed and not enough catalyst and too much catalyst and not enough feed. This is easily pictured by viewing the catalyst density profile across the Riser.



Elevation	Density (LB/FT³)	
	Low Steam	High Steam
6'-9"	1.7	1.8
9'-0"	1.4	2.1
10'-10"	2.4	2.7
12'-6"	4.7	4.2
13'-9"	7.1	5.7
15'-3"	10	8.3

Figure: Density profile of

a FCC riser

6.5. Poor Stripper Operation

Stripping efficiency is often a victim of the increased feed rates and resulting catalyst circulation over the years. As catalyst circulation (and flux through the Stripper) increases, without commensurate rising of Stripper parameters (both mechanical hardware and operations), countercurrent contacting of the spent catalyst with steam becomes less efficient, and more of the 146nstrapped hydrocarbons are carried into the Regenerator.

These result in:

a) Loss of yield and conversion
b) Consumption of air for the combustion of 146nstrapped hydrocarbons instead of hard, catalytic coke
c) High Regenerator temperatures or increased after burn.

Inefficient stripping could involve hardware design limitations (Stripper diameter, trays, and open area) as well as operating limitation (inadequate or mal-distributed stripping steam).Note that, at high enough catalyst flux rates, increasing stripping steam has not benefit as much of the added steam is simply carried down into the Regenerator with the catalyst.

6. 5.1 Analysis of Stripper Operation

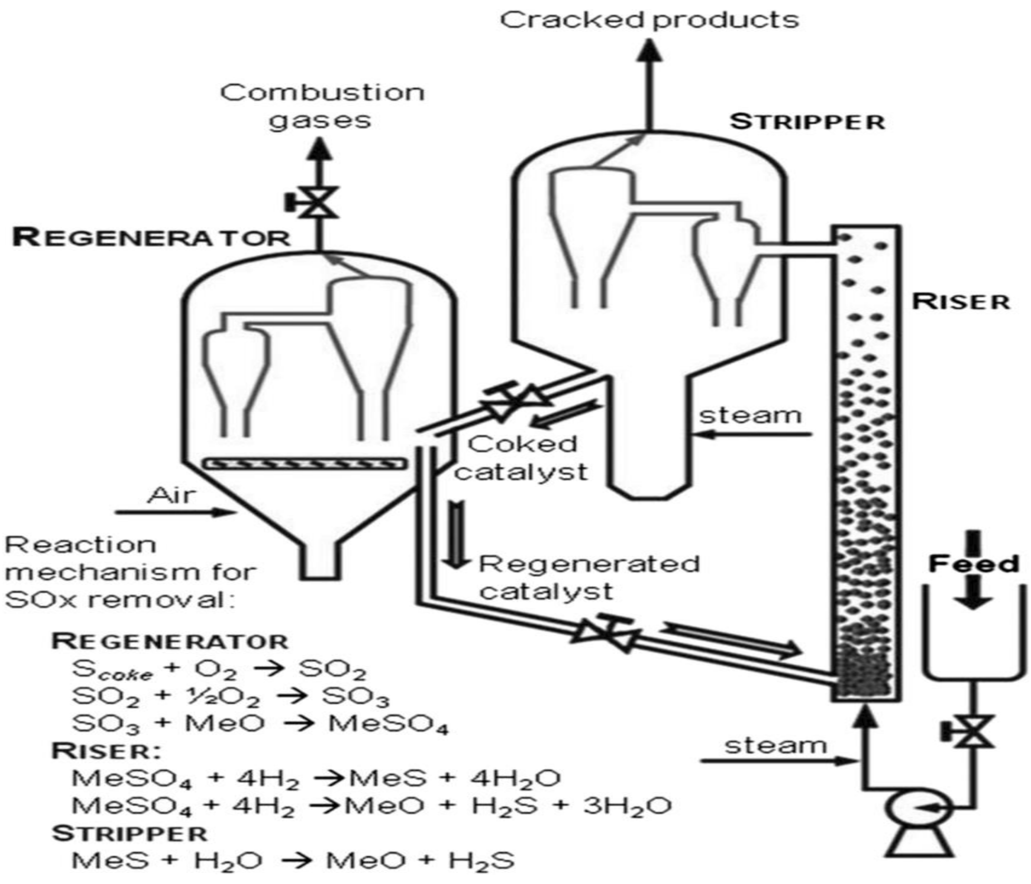
A number of methods exist for determining whether a Stripper is adequately designed and operated based on current needs. These are presented in the approximate order of complexity and required effort.

Basic calculations: Based on experience with various designs of Strippers, formulae and correlations exists which allow the expert to use simple calculations involving hardware parameters (i.e., Stripper type, diameter, length, area, volume, restricted area; tray design, tray types and numbers and gaps and holes; steam distributor design parameters, quantity, and location) and operating parameters (i.e., catalyst circulation, pressure and temperature, steam rate, Stripper catalyst density, superficial and restricted flux, steam velocities and residence times) to estimate the extent to which the Stripper might be overloaded

6.5.2 Improving Stripper Operation

Depending on matters of schedule and urgency and capital restrictions, and the understanding of the root cause of the problem, a number of options exist. These are listed below in approximate order of complexity.

a) Reduce catalyst flux:This is trivial as, one assumes, the increase in catalyst flux due to higher FCC feed rates is the essence of the problem.
b) Increase steam rate: If additional steam is available, it should be increased, at Least as a test to optimize steam-to-catalyst rates.
c) Modify or add steam rings or nozzles: In order to improve distribution of steam Throughout the Stripper (vertical as well as radial).
d) Modify Stripper internals: Designs have been developed for increasing the Stripper open area by reducing tray “footprint” while at the same time improving steam/catalyst contact efficiency.
e) Increase Stripper diameter:This most drastic measure requires, in practice, the design and construction of a wholly new Stripper.



Schematic representation of fcc unit with reactions.

6.6. Poor Regeneration:

The “ideal” Regenerator, at least in the dense phase , can be conceptualizeed to be very much like a classical CFSTR (albeit a two-phase one) with spent/stripped catalyst and air mixed perfectly, with no mixing and separation zones, and no boundary layers resulting in temperature or concentration/density variations. To the extent that the dense phase parts from this ideal (i.e.,imperfect distribution of catalyst, air, and temperature), problems result. The improper mixing can be radial or tangential due to mis-design of, malfunction of, or damage to systems used to introduce the catalyst and air into the Regenerator and those for removing the regenerated catalyst and flue gas out: slide valves and nozzles, showerhead, air rings, sparger pipe.

Another mode of improper mixing involves the vertical jetting of air beyond the dense bed surface into the dilute phase. This results in sub-optimal use of oxygen, not to mention high catalyst losses into the flue gas. The distribution and mixing of catalyst and air can also be adversely impacted by spent catalyst particle size distribution. Spent catalyst with too much “fines” is entrapped more easily into air jets and carried to the dilute phase where as catalyst with not enough fines might result in significant dead or slow spots where mixing inadequate.

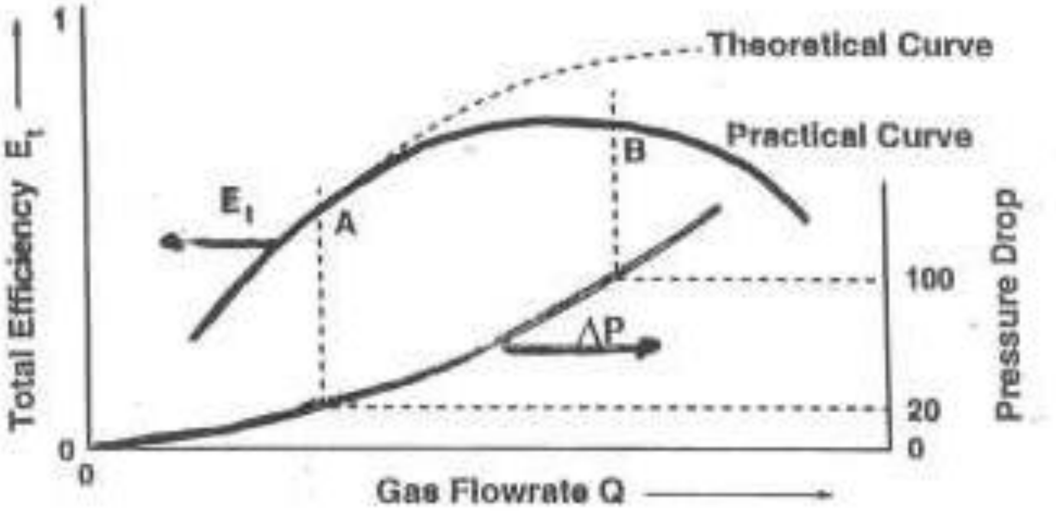
An additional key source of poor regeneration is inadequate oxygen. This problem becomes common as the FCC is pushed beyond original design limits with higher feed rates and lower quality (heavier) feeds.Note that poor regeneration is separate from the reduction of E-cat activity caused by deposition of metals such as vanadium, nickel, iron, etc. Whereas coke on catalyst is, almost by definition, reversible through regeneration with oxygen, the latter, metals on catalyst can be either irreversible (and must be withdrawn from the system and discarded) or require processing, again outside the FCC system, using novel technologies.

6.7. Poor Cyclone Performance

Problems with cyclones in the FCC Regenerator or Reactor systems present a significant issue with refiners due to the impact on operational stability, variable costs and profitability and potential environmental effects. The most common manifestation is increased catalyst losses from the Regenerator into the Flue Gas system or from the Reactor into the Main Fractionator. Two Stage Cyclone systems that are properly designed and operated within normally acceptable ranges of velocity and mass flux can attain overall catalyst fines recoveries above 99.99%. However, when the conditions fall outside of normal criteria, cyclone performance can rapidly deteriorate.

Among the many causes of poor recoveries are eroded barrels, hopper cones and diplegs, internal refractory or metallurgical component damage, excessive catalyst loading, high inlet or outlet velocities, coke or refractory plugged diplegs and immovable dipleg outlet flappers. “Soft” fresh catalyst or catalyst with an abnormal particle size distribution can also contribute to higher losses.

Many of these effects can be corrected by applying the proper design standards in relation to the typical operation of the FCC. However, in some instances, increased catalyst losses can occur although all the design and operational parameters are within acceptable ranges. In these cases, it is important to be able to identify that losses have increased and enact the appropriate response.



Cyclone efficiency curve : Efficiency vs. Vapor flow rate and pressure drop.

6.8. Maintenance of Catalyst Properties

Maintenance of catalyst activity is critical to optimized FCC operation and profitability. The loss of yield, conversion and selectivity can be acute if catalyst contaminants and microactivity are not sustained at appropriate levels. Effective catalyst management must be practiced to minimize delta coke and dry gas yields.

Increasing concentrations of these additive contaminants function to shift the reaction mechanisms to those less desirable to the FCC processing goals.

The main negative aspects of nickel accumulation are:

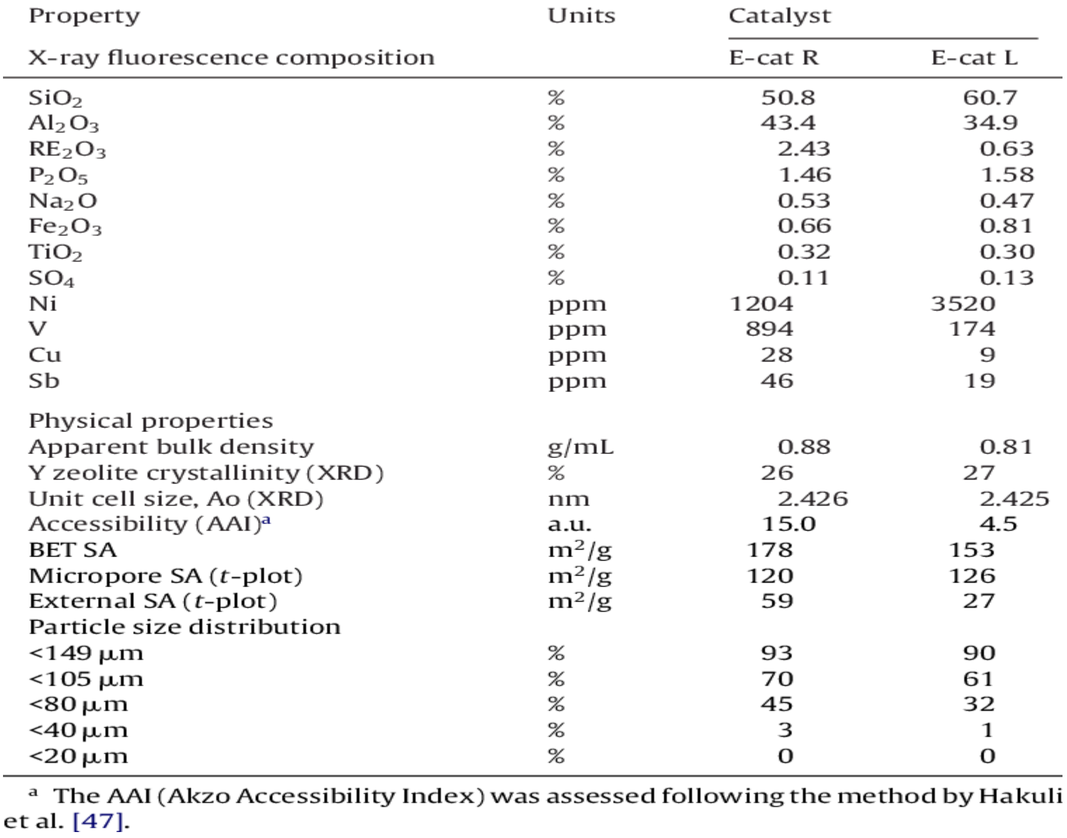
a. Deposits on catalyst surface with minimal migration
b. Catalyzes undesirable dehydrogenation reactions and coke
c. Slightly decreases equilibrium MAT
d. Increases delta coke and decreases catalyst circulation rate and cat-to-oil ratio, leading to lower conversion, lower gasoline yield, lower gasoline selectivity, and lower total liquid yield. Fuel gas yield and olefinicity of liquid products will be increased.

The main negative aspects of vanadium accumulation are:

a. Migrates to the active zeolite components and destroys them
b. Decreases equilibrium MAT
c. Slightly catalyzes undesirable dehydrogenation reactions and coke
d. Decreased MAT leads to lower gasoline and LPG liquid yields, lower gasoline selectivity and lower total liquid yield.

Overall, high catalyst contaminant concentrations are always unfavorable as they result in gasoline and LPG product yield decreases with corresponding increases in Light Cycle Oil (LCO), Slurry, Dry Gas and Coke. Whether by reaction mechanism or zeolite destruction, delta coke increases with a resultant increase in Regenerator temperatures, loss of catalyst circulation rate and loss of cat-to-oil. Stripper operation may also be affected due to increased heavier, hydrocarbon under carry.

Figure: Properties of catalyst



Property	Units	Catalyst	
X-ray fluorescence composition		E-cat R	E-cat L
SiO_2	%	50.8	60.7
Al_2O_3	%	43.4	34.9
RE_2O_3	%	2.43	0.63
P_2O_5	%	1.46	1.58
Na_2O	%	0.53	0.47
Fe_2O_3	%	0.66	0.81
TiO_2	%	0.32	0.30
SO_4	%	0.11	0.13
Ni	ppm	1204	3520
V	ppm	894	174
Cu	ppm	28	9
Sb	ppm	46	19
Physical properties			
Apparent bulk density	g/mL	0.88	0.81
Y zeolite crystallinity (XRD)	%	26	27
Unit cell size, Ao (XRD)	nm	2.426	2.425
Accessibility (AAI)[a]	a.u.	15.0	4.5
BET SA	m^2/g	178	153
Micropore SA (t-plot)	m^2/g	120	126
External SA (t-plot)	m^2/g	59	27
Particle size distribution			
<149 μm	%	93	90
<105 μm	%	70	61
<80 μm	%	45	32
<40 μm	%	3	1
<20 μm	%	0	0

[a] The AAI (Akzo Accessibility Index) was assessed following the method by Hakuli et al. [47].

CHAPTER-7

7.1. PLANT UPSET CONDITIONS, STABILIZATION AND AVOIDING UP

The effects of various operating and design conditions were to be tested and its effects on the naptha yield ,coke yield and total conversion were to be noted.Conditions for the given problem .preheat crudies passed through a separator in to a vapour and liquid.The flash evaporator flashes the liquid crude oil in to a vapour+ liquid mixture of about 50%.Both the components passes then to the through mixer and then to the fractionating column.Atmospheric gas oil to the FCC unit for cracking.

- Fresh Feed Quality
- Reactor Operation
- Fractionator Operation
- FGD-II Operation
- Fresh Catalyst Quality
- Regenerator Operation
- Gas Compression Unit

The problems if not properly encountered may lead to unit emergencies or poorer product yield, loss of equipment, environmental issues etc.

7.2. FRESH FEED QUALITY

Feed quality and its effect on the performance of the fluid catalytic cracking unit as a case study. The important feed qualities used are the hydrocarbon content and a hydrotreated feed. Data on the feed properties used in PHRC were collected and a product mass balance was carried out on the fluid catalytic cracking unit. Conversion and gasoline yield of the unit were found to be 73.43 vol% and 52.07 vol% respectively. On comparison with cracking of aromatic feed, from literature, with 61.3 vol% conversion and 45.64 vol% gasoline yields, the feed is said to be paraffinic because of its higher conversion and gasoline yield. On comparison with that collected for hydrotreated feed, 80.62 vol% conversion and 63.9 vol% gasoline yield, it isconcluded that feed hydrotreating increases conversion and gasoline yield by a significant ammount.

EFFECTS:

- High H_2 making
- overloaded wet gas compressers
- Increase in height in light end production
- Leak in preheats
- Feed type change
- Loss of fines

- **STABILIZATION**
- Lower metals in feed
- Feed segretion
- Reduce slop cut usage

- Minimize catalyst losses on tank heaters
- Change of fine catalyst

7.3. REACTOR OPERATION PROBLEMS:

- Feed nozzles plugging causes the loss of stream purge during feed outage resulting in catalyst backup plugging by refractory and catalyst during startup.
- It causes the Erosion and/or corrosion.
- Pressure surges,Condensation of reactor products in the overhead line, Insufficient stripping steam,Poor catalyst/steam contacting Catalyst properties,Low reactor temperature Inaccurate stripping steam flow controller.

7.4. REGENERATOR OPERATIONS PROBLEMS

- Upset conditions for this cyclone failure,plugged diplegs,afterburning ,grid hole plugging and or erosion are the main problems.
- Excessive temperatures allowable stress exceded,spilled refactory forming partial and eventually total plugging, insufficient coke production
- Areas of defluidized catalyst and low bed stability.

EFFECTS

- High catalyst content in fractional bottoms
- Catalyst plugging in slury lines/slurry pumps suction strainer
- Frequent slurry pump loss of suction
- Loss of catalyst fines

STABILIZATION:

- Minimize reactor velocity
- check stripping steam rate
- consider lower circulation rate
- carefull detailed check of the unit to prevent refactory spilling

7.5. FRACTIONATOR OPERATING PROBLEMS

- Excessive corrosion in overhead system
- poor fractionation between side streams
- This causes the inadequate wash water flow rate
- High contaminant cocncentration in sour water stripper bottom

EFFECTS:

- Scale from sample outlet ,high corrosion rate measured by probes ,High overlap between stream distillations,shift in yield pattern
- increase in one side stream volume at constants operating conditions

STABILIZATION:

- Increase wash water and corrosion inhibitor rates
- Adust pump around heat duties consider equipement changes

7.6. FLUE GAS DESULFRIZATION UNIT PROBLEMS:

- Scrubber quench water failure
- sudden loss of quenchspray tower(14-c1001) level
- causes malfunction of quench tower recicuraltion pumps
- upset of FCCU

EFFECTS:

- Sudden rise of scrubber quench zone
- Damage to scrubber and internals mechanical integrity if failure of quench liquid further persist

STABILIZATION:

- Its quick identification of source of the problem.
- if required make_up control valve by pass needs to be operated.
- if level is further dropping,emergency dilution SDV has to be operated for making up the level till the make_water flow is normalised.

7.7. AVOIDING DEVIATIONS AND PLANT UPSETS:

Upset conditions do occur in plant operations due to any unforesen activity or emergency occurrence either in the plant or outside the battery limits.Avoiding the conditions for the upset at the base level is utmost important forsmooth operation of the unit.some incidents and upsets can be avoided by timely action and precautions.proper checking in that field ,following DRJs, close monitoring of process parameters at DCS and ensuring that all systems and procedure are followed is the only way to avoid upsets.Monitoring of pumps and their auxuliary systems at regular intervals will help in taking corrective action.The results in increased pump life and thus lesser upset due to pump failures.Gauge glasses need to be checked and tailed with Lts which helps in identification of LT malfunction.Ensuring healthiness of steam tracings and EHTs will minimize the risk of plugged hydrocardon lines.PM checks of vritical motors and instruments shall be taken up.

7.8. SOME OF THE PRECAUTIONS FOR AVOIDING THE PLANT UPSET PROBLEMS:

1. Catalyst loss from reactor/regenerator cyclones:

I. Maintain minimum cyclone inlet velocities as per design (> 18 m/sec).
II. System pressure can be slightly reduced at lower throughputs to improve the velocities.
III. Lift steam in reactor to be increased.
IV. Ensure proper amount of fines are present in catalyst inventory.

2. Plugging of feed nozzles:

I .Close monitoring of DP across feed filters.
II. Purging feed nozzles with high steam rates during feed failure and prior to catalyst loading during start up.

3. **After burning / Behind-burning in regenerator :**

I.Air to regenerator, Reactor temperature and strippingsteam in reactor are to be adjusted depending on the feed rate and quality of feed.

4. **Slurry circuit coking:**

I. Maintain minimum fluidization velocities of slurry streams.
II. Monitor slurry catalyst content and take necessary corrective action.
III.Cleaning of slurry pump suction strainers on regular basis.

5. **Stabilizer pressure fluctuation:**

I. Avoid abnormal increase in reflux flow.
II. Ensuring steady reboiler outlet temperatures.
II. Avoid slippage of C1 and C2s into stabilizer feed by ensuring stripper re-boiling.

CHAPTER-8

FUTURE ADVANCEMENTS/DEVELOPMENTS IN FCC.

1. Sulphur-reducing chemistry that preserves gasoline yield and minimises octane loss while avoiding cut point penalties can help meet clean fuel specifications where post treatment and additive systems may not be adequate.
2. One option being considered is the use of a catalytic sulphur reduction additive, with reductions of up to 30% reported. Another option is to reduce the end point of the FCC naphtha.
3. Since most of the sulphur is concentrated in the “back end” of the gasoline, this can effectively reduce the sulphur level, but with the penalty of reduced gasoline yield.
4. Engelhard has recently commercialised a new FCC catalyst technology,Naphtha Max, which has demonstrated significant improvements in conversion and gasoline yield in multiple commercial applications.
5. The technology features Engelhard’s Pyrochem-Plus zeolite with a new matrix technology known as distributed matrix structures (DMS) to achieve optimised.
6. along with the high activity needed for today’s short contact time (SCT) cracking applications.
7. The FCC regenerator design has a direct impact on the effectiveness of in-situ reduction of NOX to N2, and hence reduction of the final NOX emissions in the flue gas.
8. The new Shell Global Solutions’ low NOX regenerator technology involves an improved process consisting of the strategic design of catalyst and air distributions, which enables the unit to operate in both full and partial combustion modes with low NOX emission.
9. FCC units are less common in EMCA because those regions have high demand for diesel and kerosene, which can be satisfied.As of 2006, FCC units were in operation at 400 petroleum refineries worldwide, and about one-third of the crude oil refined in those refineries is processed in an FCC to produce high-octane gasoline.During 2007, the FCC units in the US processed a total of 5,300,000 barres (840,000m^3) of feedstock per day,and FCC units worldwide processed about twice that amount.
10. In the US, fluid catalytic cracking is more common because the demand for gasoline is higher.
11. The present paper will focus on recent progress in FCC catalyst and process technology and will analyze the driving forces for such improvements.
12. The direct response to environmental regulations will become obvious.Another target of new developments has been to process heavier crude sources with higher levels of contaminants.

13. Short contact time(SCT) cracking will also be discussed,reflecting significant

developments in the reator hardware.

number of advantages:

1. Improved accuracy and repeatability
2. More automation – less operator-induced variability
3. Heavier feed capability
4 .Better heat transfer/lower temperature drop
5.Better feed/catalyst contacting
6.Uniform coke deposition–no bed gradients
7.Improved stripping in fluid bed

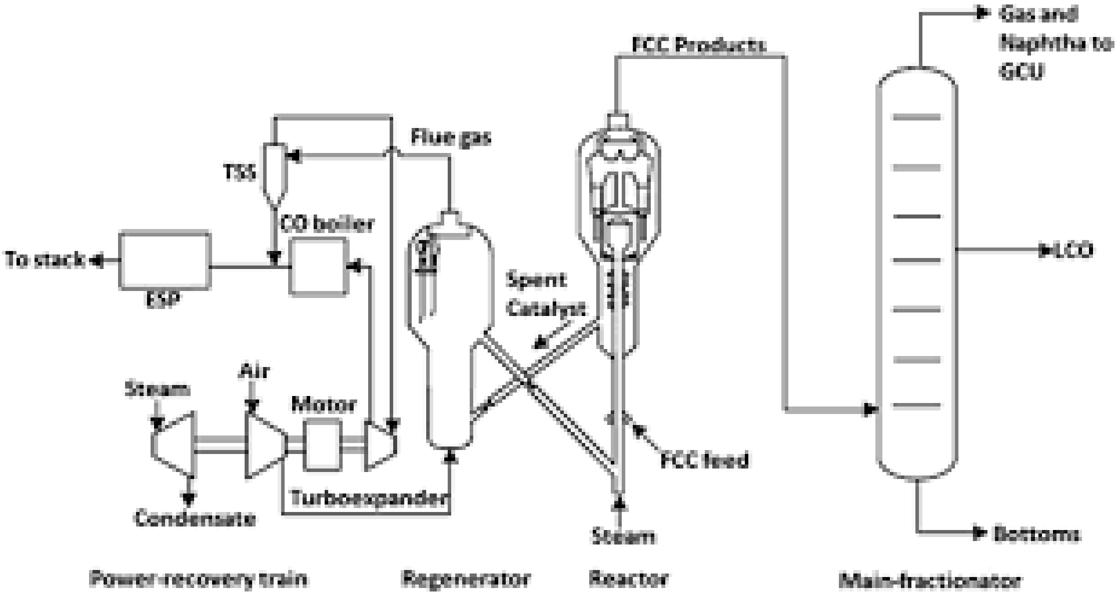


fig:Recent developoments in fcc process and catalyst:

CHAPTER-9

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