

Modeling and Simulation of Fluidized Catalytic Cracking Riser Reactor Using Pseudo Reaction Kinetics : A Review

P.K.Yadav^{1*}, Rajeev Kumar Garg²

¹ Research Scholar, IKG Punjab Technical University, Kapurthala ,Punjab(India)

²Department Of Chemical Engg.,Shaheed Bhagat Singh State Technical Campus,Ferozpur(Punjab-India)

¹pkyadav75@yahoo.co.in

²rkg_garg@rediffmail.com

Abstract: The fluid catalytic cracking(FCC) is one of the key process unit in modern oil refining . FCC process converts heavy distillates like gas oil or residue to gasoline and middle distillates using cracking catalyst.FCC consist of two interconnected gas –solid fluidized bed reactor: riser and regenerator .FCC process is very complex and require good understanding of many factors such as kinetic models, reaction kinetic, fluid dynamics, feed and catalyst effects, temperature, residence time in the riser, catalyst deactivation and its regeneration. The performance of the reactor are measured in term of yield of gasoline and light gases, feed conversion, coke yield etc. A number of operating variables can be optimized for economical operation of FCC unit.

Key words: FCC; Cracking Catalyst; Riser and Regenerator; Kinetic model; Gasoline yield.

I. Introduction

Fluid catalytic cracking unit (FCCU) is used in most refineries to convert heavy gas oils feedstocks [1-3] into more valuable hydrocarbon products such as gasoline, Liquefied Petroleum Gas(LPG) along with dry gases and coke inside a riser reactor using cracking catalyst[4].The feeds used in riser include molecules with carbon number (C_{30s} to C_{60s}) molecules mainly in form of cycloparaffins (naphthenes),aromatics and olefins. The product yield of catalytic cracking depend upon feedstock quality. Main product classification from cracking include gasoline (C_{5s} - C_{12s}),Liquefied petroleum gas LPG(C_{3s} & C_{4s}) and light cycle oil (LCO: C_{11} through C_{18}).Some by product such as dry gases ($H_2,C_1,C_{2's}$),sour gas(H_2S),heavy cycle oil (HCO: C_{19s+}) considered as unconverted feedstock and solid Coke[5-7]. FCC contributes to the bottom up gradation far more than other processes. The fluid catalytic cracking unit is the largest producer of gasoline and light ends in the refinery[8].The unit can handle a wide range feedstock and can be operated in maximization modes for gasoline, LPG or diesel without involving any hardware changes as per demand in different seasons .FCCU can rightly be called as ‘Heart of a Refinery’ and the economic benefit of refinery can be increased by proper control & optimization strategies. The control of FCC processes have been known as challenging problems due to complicated and little known hydrodynamics, complex kinetics, strong interactions between the reactor and generator and many operating constraints.

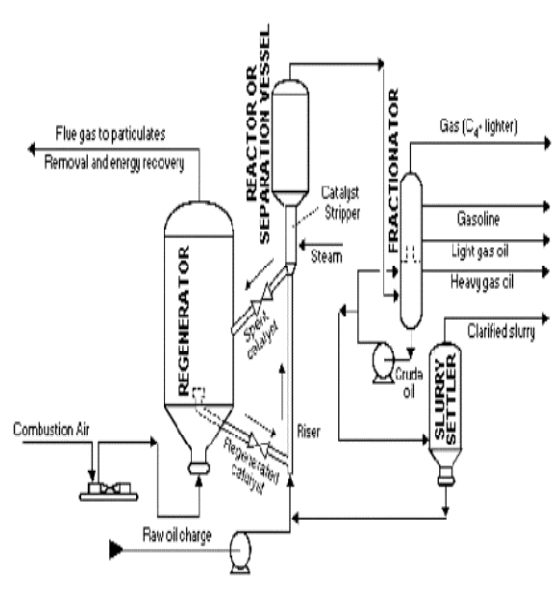


Fig.1. Fluidized Catalytic Cracking Unit

The FCC units are consist of two major operating parts, the reactor riser & the regenerator. The riser where almost all the endothermic cracking reaction of the hydrocarbon feed and coke deposition takes place. The regenerator reactivates the catalyst by burning the accumulated coke on the catalyst in the riser reactor by use of air. The regenerator process provide heat required for endothermic cracking process.[9] The feed preheated to a temperature of 177-327 ° C in a furnace & this feed is injected into the bottom of the riser along with a small amount of steam (0.5-3 wt.% of feed) which leads to good atomization & reduces coke formation. The entering feed is vaporized upon contacting the hot catalyst flowing from the regenerator. A residence time of 2-5 seconds of catalyst & hydrocarbons vapors is preferred in the riser. The riser top temperature in range of 477-547 ° C is controlled by regulating the flow of hot regenerated catalyst to the riser[4]. The reactor temperature is kept constant by adjusting the catalyst/oil ratio or else by keeping the outlet temperature from the preheating surface at a suitable level. The basic tubular design of the riser reactor is common to all existing units. The main difference in the designs is the way the catalyst circulation between the regenerator and reactor is controlled. The disengaging section of the riser used to separate the catalyst particles from the vapors. The product vapors from the disengaging section enters a main distillation column where vapor products are separated into various boiling point fractions. The spent catalyst is separated from the vapor in the reactor cyclone and falls into the stripping section where the hydrocarbons remaining on the surface are removed by the stripping steam. The stripped spent catalyst is recycled through a catalyst transport line to the regenerator.

In the regenerator which is operated in fluidization regime, the coke is burnt off the catalyst surface by the hot air blown into the bed. Carbon can be converted to CO or CO₂. The combustion reaction serves to reactivate the catalyst and to maintain the bed hot enough to supply the heat required for the vaporization and cracking reaction of the feed in the reactor [10]. The typical catalyst residence time in the regenerator ranges from 5-15 min. Typical regenerator temperature range between 675-760 ° C. Combustion products and entrained catalyst are conveyed upward, out of dense fluidized bed, into a dilute phase zone where cyclones separate the catalyst, which is returned to the bed. A FCC regenerator operate in turbulent fluidization and there is considerable carryover of the catalyst to the cyclones. In fact the whole bed circulate through the cyclone every 5 minutes. The regenerated catalysts flows continuously into the riser bottom through catalyst transport line. Cracking units are equipped with one or more coolers to remove excess heat. In modern units the catalyst circulation is controlled using a slide valve in the catalyst transport line feeding the hot catalyst to the reactor, while in the older units the catalyst circulation rate could only be adjusted over a narrow range by the pressure balance[11].

II. Literature Review

Brief review on the work done on the modeling of FCC riser is presented in this section. In the refinery FCCU play a very important role for economic point of view. In the modeling and operation of FCCU, the FCC riser & regenerator requires different modeling schemes and parameters based on reactions in continuous mixtures, structure oriented lumping and single –events cracking in riser modeling and regenerator hydrodynamics, combustion reaction kinetic and temperature control in regenerator modeling. The application of these model is difficult because of the analytical complexities and computational limitations. Modeling of the riser reactor of an FCC unit is quite complex because of the presence of all three phases(Solids, Liquid, and vapor) inside the reactor, involvement of physical and chemical steps, and its strong interaction with the regenerator. A lumped kinetic model which utilized pure components cracking data for the catalytic cracking of oil mixtures by considering the petroleum feedstock into a number of pseudo components is generally considered. The number of operating parameters like feed temperature of oil, catalyst circulation rate, number of feed nozzle, temperature distribution with in reactor, pressure drop, catalyst hold up, deactivation of catalyst are important for riser and regenerator combustion reaction kinetics, temperature control in regenerator are important for regenerator. In modeling there is phase flow problem, which need to be considered. The flow pattern, size distribution of particles, solid properties, bed diameter and operational conditions are some parameters that need to be considered in dynamic behavior of riser.

2.1 Kinetic Model

In the first three lump kinetic model proposed by Weekman & Nace (1970), the charge stock and products divided into three components namely, the original feedstock, the gasoline, and the remaining C₄'s (dry gas and coke).

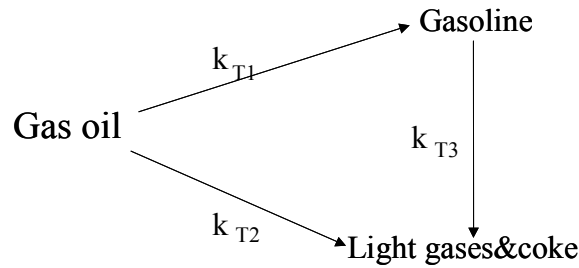


Fig. 2. Three Lump kinetic Model

Yen and Woei (1988) and Lee et al (1989) developed a four lump kinetic model by dividing the light gas plus coke lump into two different lumps C₁-C₄ gas and coke. The yield of C₁-C₄ gas increases with increasing reaction temperature at the expense of the yield of gasoline and coke. In the advanced models of FCC riser model, three lump or four lump kinetic schemes are being generally used to avoid the mathematical complexities and load of computation that will be there if more number of lumps are considered[11].

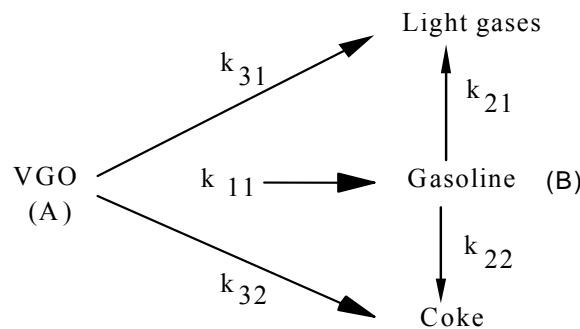


Fig.3. Four Lump kinetic Model

The rates of gasoline formation and gasoline consumption were described as,

a-) gas oil consumption rate

$$r_{go} = -(k_{11} + k_{32} + k_{31})\phi C_{go}^2 \dots(i)$$

b-)gasoline formation rate $r_g = k_{11}\phi C_{go}^2 - \phi(k_{21} + k_{22})C_g \dots(ii)$

c-)light gases formation rate $r_{lg} = k_{31}\phi C_{go}^2 + k_{21}\phi C_g \dots(iii)$

d-)coke formation rate $r_c = k_{32}\phi C_{go}^2 + k_{22}\phi C_g \dots(iv)$

k=rate constant , ϕ =decay of catalyst activity, Ci=Molar concentration of Lumps

More detailed kinetic schemes were presented by various researchers. Corella and Frances(1991) developed a 5-lump models ,in which the gas-oil lump was divided into its heavy and light fractions. Dupin et al (2006) simplified the 5-lump model of Corella and Frances. Another 5-lump model was developed by Larocca et al (1990) in which the 3-lump model of Weekman was modified by splitting the gas oil lump into aromatic, paraffinic and naphthenic lumps. Ancheyta et al (1999) followed a different approach in their 5-lump models development, in which they considered the gas oil as one lump, but divided the gas lump into two lumps liquefied product gas and dry gas. Hagelberg et al (2002) expanded the 5-lump model of Anchetya –Jurez et al (1999) to an 8-lump model by dividing the gasoline fractions into paraffins, olefins, naphthenes and aromatics[2].

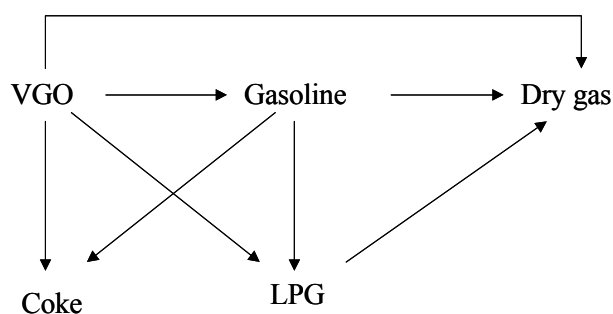


Fig.4. The five lump model

A new reaction kinetic model was developed by Jacob et al by dividing the feed and products into a 10- lump reaction schemes including paraffins, naphthenes, aromatic rings and aromatic substituent groups in light & heavy fuel oil fractions. The new model considered the feed properties in addition to the boiling range. The advantage of this lumped model is that the composition of lumps can be measured with various experimental techniques. The rate constant which arises from the use of model are less sensitive to changes in feed and process conditions.

Table:1 Components of 10 lump kinetic model proposed by Jacob et al

Initial Components in Feed	Cracked Product	
Heavy paraffinic (Ph)	Light Paraffinic (Pl)	Light Aromatic(AI)
Heavy naphthenic (Nh)	Light Naphthenic(NI)	Gasoline(G)
Heavy Aromatic substituent's (CAh)	Light Aromatic substituent's (CAI)	Coke & Light gases (C)
Heavy Aromatic (Ah)		

With increase in number of lump, the requirement of cracking constant increases rapidly. These lumps are to be characterized as an indicator of the feedstock composition.

A relatively new method, called structure-oriented lumping(SOL), for describing the composition, reactions and properties of complex hydrocarbon mixtures has been developed by Quann and Jaffe(1992). The model tracks different compositional and structural attributes of a molecules(number of aromatic rings, number of nitrogen substitutes, sulfur substituent's, etc.) in a vector format[8].

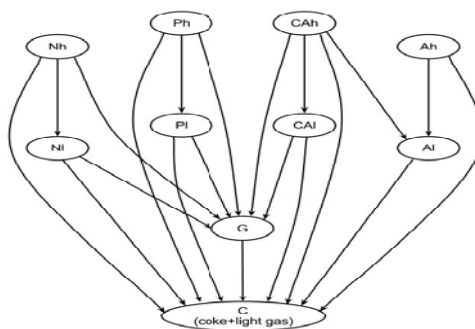


Fig.5. 10 lump kinetic model proposed by Jacob et al.

Pitault at al., proposed a model where lumps were classified by chemical family (e.g. paraffins, olefins, naphthenes, and aromatics) and molecular weight or boiling range. This approach was applied to the light gases, gasoline, LCO and feedstock cuts. It has to be emphasized that this model has the flexibility of being able to include important reactions like coking and hydrogen transfer.

Many researchers have developed complex reaction schemes based on basic chemical principles that involves thousands of chemical species. These can be classified into mechanistic models and pathway models. Mechanistic model track the chemical intermediates, such as ions and free radicals, that occur in the catalytic FCC process. Froment and co-workers proposed single event models involving a detailed description of VGO catalytic cracking. This model considered, between other factors a mechanism including carbenium ions. However, the second generation of FCC models had many drawbacks, the extensive computational calculations due to large size and reaction complexity.

N.V.Dewachtere et al (1997)[12] developed a riser model that accounts for the detailed, quantitative description of the cracking kinetics, based on elementary steps and single event coefficients and for the detailed hydrodynamics. A three –dimensional ,two –phase mathematical model for the simulation of the flow and the heat transfer in a riser reactor was developed.

The reactor model can predict the important features of a riser reactor, including pressure drop, catalyst hold up, interphase slip velocity, acceleration of the catalyst in the inlet zone and the temperature distribution of both phases over the reactor. In the research it was predicted that secondary over cracking reactions in the inlet zone ,resulting in high product compressor energy consumption were reduced by increasing the number of feed injection nozzles and that the most uniform temperature profile in the inlet zone was obtained for a feed nozzle inclination of 45.

In –Su-Han et al (2000)[2] developed a detailed dynamic model of a typical fluidized catalytic cracking (FCC) unit that consist of the reactor, regenerator, catalyst transport lines and several auxiliary unit. Special attention has been paid to the reactor riser to predict the velocity distributions of the catalyst and gas phases, the molar concentrations of 10 –lump species and the temperature profile by utilizing momentum, mass and energy balances. The regenerator model described the behavior for two regime (dense bed and freeboard) , and two phase (emulsion and bubble) .

J. Ancheyta-Juarez et al (2001)[7] studied on experimental evaluation of Vacuum gas oil-light cycle oil blends as feedstock. The effect of up to 12.8 vol % of LCO in the typical FCC feed was studied in a micro activity unit. A decrease in conversion of gasoline and LPG yields was observed when LCO is added to the conventional FCC feedstock. The opposite behavior was observed for dry gas and coke yields. Gasoline RON increased as the LCO in the feed was also increased. This was attributed to the high aromatic contents in the feeds containing LCO .The study has found an optimum content of LCO in the FCC was to be 7 vol % ,which maximizes both gasoline production and gasoline RON.

G.M.Bollas and I.A.Vasalos et al (2004)[5] in their research studies on bulk molecular characterization approach for the simulation of FCC feedstocks. In their study they had presented a bulk molecular characterization approach for the simulation of FCC feedstock regarding their relative potential to enhance catalytic cracking and coke formation .The final models for the prediction of crackability and coking tendency incorporated the specific gravity, the refractive index, and the TBP distillation of the feedstocks, along with the sulfur ,nitrogen, and carbon contents. These properties were combined properly to derive functional groups that accurately characterize the behavior of FCC feedstocks at cracking conditions. The impact of the average hydrocarbons structure on catalytic cracking was expressed via the functional group called crackable carbon atoms which describes the effects of the average molecular structure and size of the feed stock on its crack ability. In their study they have found that the aromaticity and the size of the average hydrocarbon were shown to dominantly impact the crack ability, whereas the total nitrogen content was verified to be the major catalyst inhibitor.

Raj Kumar Gupta et al (2007)[13] developed a kinetic model for the fluid catalytic riser. The model incorporated a more realistic kinetic scheme for the cracking reactions, and a new correlation to evaluate Arrhenius type reaction rate constants. The model is capable of predicting overall conversion, products yields, temperature , and catalyst activity along the riser height. These predictions of the FCC riser reactor model are dependent on the values of cracking reactions rate constant for different characteristics of the feedstock, type of catalyst, activity of catalyst and operating parameters.

G.M Bollas et al (2007)[2] studied the five –lump kinetic model with selective catalyst deactivation for the prediction of the product selectivity in the fluid catalytic cracking process .Four scenarios were examined in which catalyst deactivation effect was considered on lumping models for simulation of FCC process. The study showed that the non –selective catalyst deactivation model performs satisfactorily. The only case in which a substantial improvement was made to the model was when a product oriented selective catalyst deactivation model was used.

Mehran Heydari et al (2010)[14] studied the seven –lump kinetic model in the fluid catalytic cracking unit. The simulation result predict that much of the gas oil conversion occurred in the first 5m of the riser, due to high catalyst concentration, high activity, highest concentration of gas oil vapor at the base. The reaction rate of gas oil to products were greatest at the bottom of riser, contributing to rapid conversion. The results show that ,because of high conversion in the initial section of riser ,with increasing of inlet temperatures, gasoline yield increased, but after the gasoline yield passed through a maximum value, because of coke deposition, decreasing of the gasoline yield occurred . The highest yield was obtained at 1075 K and was a function of feed quality, catalyst type and reaction temperature.

Kiran Pashikanti et al (2011)[8], developed a model for a FCC unit that includes a significant implementation of the associated gas plant using Aspen HYSYS. In his work, model was given which described how to use routinely collected plant data with well known commercial software tools to present an integrated process model that include both reaction and fractionation systems. An integrated model allows users to identify opportunities to improve yields, increase profitability, and monitor the unit for predictable operation.

Prabha K.Dasila et al(2012)[15] simulated an industrial FCC unit by integrating kinetic models for the riser reactor and the regenerator. The model was used to study parametric sensitivity such as effects of feed preheat temperature, feed flow rate and air flow rate on the FCC performance.

K.K.Dagde et al.(2012)[9] proposed a model to study the operation of the riser and regenerator reactors in an industrial FCC unit . The model adopted five –lump kinetic scheme. The result indicate that catalyst to oil ratio and inter air velocity have significant effect on the performance of the riser and regenerator respectively. The optimum values of process variables obtained in the analysis for effective operation of FCC were inter –air velocity of 14 m/s, riser temperature of about 653 K, and catalyst –gas oil ratio of 3.

Babatope Olufemi et al (2013)[16] presented a one dimensional adiabatic model of FCC unit riser reactor for a numerical analysis of the process. The effect of COR and other variables were investigated by the model. The yields, temperature and pressure profiles along the height of the riser reactor were computed by using the model.

H.A Affum(2013)[17] proposed a kinetic model in combination with the tank in series model, used to predict the conversion response of an FCC riser to changes in feed temperature ,feed flow rate as well as riser diameter and height (to vary residence time).Results indicate that shortening the residence time and decreasing gas oil feed temperature may significantly reduce overall conversion.

2.2 Catalyst

Zeolite is the important ingredient of FCC catalyst. In the refinery, zeolite catalysts have been used for improvement in gasoline yield, octane number, production of cleaner fuels with enhanced performance properties. Zeolites as fluid catalytic cracking catalysts increases the formation of desired cracking products in FCC unit compare to amorphous, silica-alumina catalyst. The zeolite catalyst are more active and more selective[18]. Most commercial catalyst contains between 15 & 40 % zeolite, which is the major contributor to the catalytic activity and selectivity. FCC catalyst consist of major components like zeolite, matrix, filler ,binder & additives. The different materials in the FCC catalyst have different functions such as : Acidic functions in amorphous and crystalline matrices, physical properties and mechanical strength, metal impurities traps, combustion promoters, SO_x traps, octane booster and olifins promoters additives. An active matrix provide the cracking sites. These sites cracks larger molecules that are hindered from entering to the small pore of zeolite. It also guards the zeolite from becoming deactivated prematurely by impurities. The filler is a clay incorporated into the catalyst to dilute its activity. Kaoline is the most common clay used in FCC catalyst. The binder serves as a glue to hold the zeolite , matrix and filler together .The function of filler and binder are to provide physical properties (density, attrition resistance, particle size distribution etc) and mechanical strength. Additives are also added for enhancing cracking performance of the catalyst.

Zeolite Y is the primary zeolite component in FCC catalyst. It is relatively stable at the high temperature with in FCC unit and very efficient at catalyzing the cracking of smaller FCC feed molecules that can enter through its microspores. Recent advances in FCC catalyst have been made by modifying zeolite Y to Ultrastable Y (USY) for improved coke selectivity, higher cracking activity and greater stability through manipulation of extra framework aluminum or through the generation of mesoporosity of the zeolite crystal.(Thomas F. Degnan et al 2000)[19].The removal of alumina increased silica oxide to alumina oxide ratio(SAR),reduced unit size(UCS),lowered the sodium level of the zeolite. This changes increased the gasoline octane by raising the olefinicity. By use of more highly dealuminated USY zeolites, researchers have been able to produce distillate selective hydro cracking catalyst.

Unit Cell size(UCS), rare earth level and sodium content are the parameter that mainly governed the zeolite behavior.

UCS is a measure of aluminum sites or the total potential acidity per unit cell. A lower UCS present fewer active site per unit cell size. **Rare earth level:** The rare earth increases zeolite activity and gasoline selectivity with a loss in octane. Rare earth improves thermal and hydrothermal stability of the zeolite.

Sodium Content: Sodium content decreases the hydrothermal stability of the zeolite. It also react with the zeolite acid site to reduce the catalyst activity.

Addition of shape selective ZSM-5 additive improve gasoline octane and enhanced the production of light olefins. Antimony, Bismuth, and Tin components are used for the passivation of heavy metal. Inorganic oxide such as metal oxide(Al_2O_3, MgO) are used as SO_x reducing additives. Platinum or palladium based metal act as CO combustion promoters. Metal oxide(Al_2O_3, TiO_2), mixed oxide($BaTiO_3, CaZrO_3$) have been recommended as metal traps as additives in catalyst.

Ziyuan Liu et al (2010)[20] in his research studied on diffusion coefficient of heavy oil in FCC catalyst . The study shows that the effective diffusion coefficient decreases as the oil fraction diameter increases and the diffusion coefficients in the large -pore-size catalyst with a template is 2-3 times larger than those in the catalyst without a template. Because of the promotion of diffusion rate of heavy oil caused by the large pore size of the catalyst , the catalytic performance of macro porous catalyst is better than the conventional catalyst.

Table: 2 Physical and Chemical properties of commercial equilibrium FCC catalyst (source :Ivelina Shishkova et al ,2011)

Chemical Composition	
Al_2O_3 % wt	40.2
Na_2O % wt	0.13
Re_2O_3 % wt	1.9
Fe % wt	0.56
V ppm	25
Ni ppm	195
Physical Properties	
Average Particle Size μm	81
Average Bulk density g/cm^3	0.88
BET Matrix surface Area m^2/gm	56
BET surface area m^2/gm	165

2.2.1 MAT Reactor:

One of the most important parameter which determine the yield, the quality of FCC products and the competitiveness of FCC unit is the catalyst type .The laboratory evaluation of fluid catalytic cracking with the use of Microactivity Test (MAT) unit ,is generally used for measuring performance characteristics of experimental and commercial catalyst samples.MAT reactor is a small fixed bed reactor far from the commercial fluid catalytic cracker (FCC) reactor.MAT reactor is commonly used for testing , evaluation of FCC catalyst. The objective of MAT reactor is also used to obtain the catalyst to oil ratio, inside the riser of industrial unit at certain operating conditions.

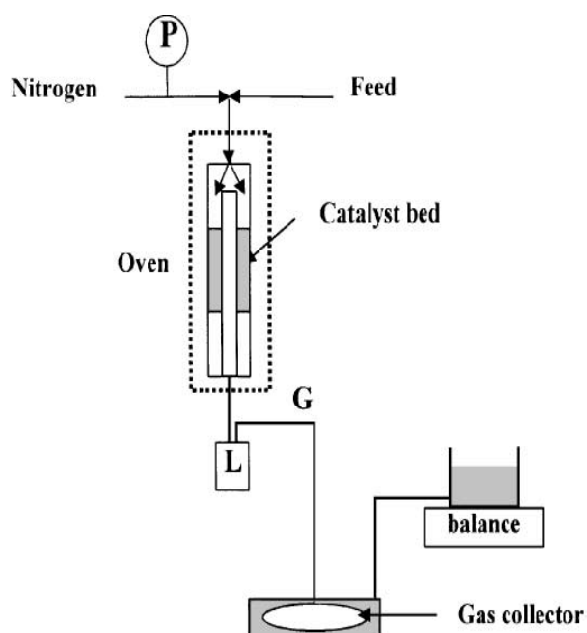


Fig. 6. Schematic diagram of MAT reactor adapted from C.Delattre et al 2001.

The Microactivity Test (MAT) unit used for the experiments are generally designed according to the ASTM D-3907 method with some modifications. It is made of a pyrex fixed bed reactor heated by a three zone furnace . The feed is injected on top of the catalyst bed for a fixed period of time over a measured amount of catalyst, using a controlled syringe pump, through an independently controlled preheat device .The products formed have to pass through the remaining part of catalyst bed exposed to secondary reactions before they leave the reactor and collected. Vapour products of the cracking are cooled to 0 °C at the reactor exit where part of them are condensed and collected in a specially designed liquid receiver. The remaining not condensed gaseous products are led to a gas collection system. Nitrogen flows are used during MAT experiments in order to drive the feed and products along the reactor and purge the injection system. The reaction products can be analyzed by gas chromatography. The catalyst activity will diminish during the whole feed injection time and product formed will therefore change with time. Catalyst to oil ratio is calculated over the whole injection time of feed. MAT catalyst to oil is commonly lower than commercial C/O used in commercial unit. During MAT experiments yield to products in the laboratory reactor is similar to that obtained in industrial unit. [21-22]

2.3 Catalyst Deactivation

FCC catalyst gets deactivated by the deposition of coke on the catalyst surface, during the cracking reactions. The catalyst deactivation also occur by deposition of nitrogen on active sites which neutralize the catalytic activity or by deposition of metals (nickel, vanadium, sodium) which can change or destroy the activity. The blocking of pore by deposition of coke or metals , reduces the mass transfer. Most of the popular theories on the deactivation are based on the time on stream concept. Various models for time dependent catalyst have been proposed for different lengths of contact time.

The overall cracking rate is affected by the catalyst activity, its effect can be incorporated by considering the deactivation model of the Weekman. In this respect, Weekman (1968) employed to describe catalyst deactivation, the following two simple relations;

a)- exponential decay law

$$\varphi = \exp(-\alpha t) \dots\dots\dots(v)$$

b) Power decay law

$$\varphi = t^{-n} \dots\dots\dots(vi)$$

where t represents catalyst time-on-stream, α (catalyst decay coefficient) and n are rate constants of the catalyst decay function.

α the catalyst decay coefficients related to the riser temperature is given by Arrhenius equation:

$$\alpha = \alpha_0 \exp(-E/RT) \dots\dots\dots(vii)$$

III. Modeling of Riser Reactor

3.1 Riser Kinetics

The feed to the FCC process consist of long chains paraffins, single & multiple ring cycloalkanes, and large aromatic compounds. The catalytic cracking begins with the formations of carbenium ions by the interaction of olefin molecules with the acidic site on the catalyst followed by the beta scission of the carbenium ion. In the beta scission reaction ,the β bond of the carbenium ion break to form an olefin and a new carbeniumion .The carbanium ion formed by beta scission can undergo further cracking reaction. The olefin can also be cracked further after being converted to a carbenium ion through hydrogen addition. Large hydrocarbon molecules can be cracked repeatedly producing successively smaller hydrocarbons, making cracking rates become slower. Less reactive ion can transfer its charge to a larger ,more reactive molecule by hydrogen transfer that makes the cracking reaction of larger molecules still faster.

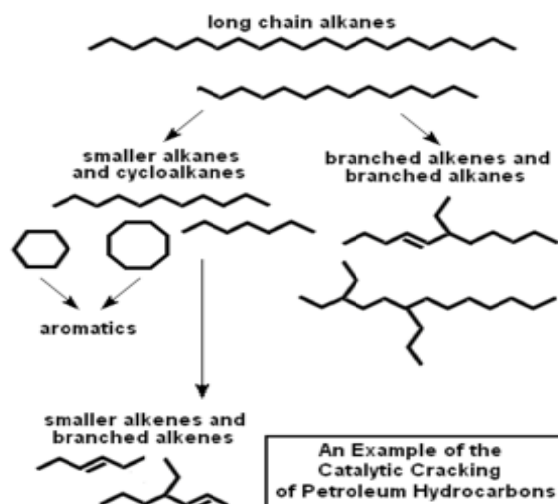


Fig. 7. Catalytic cracking of petroleum hydrocarbons

Primary reactions involved in catalytic cracking are:

- Paraffins are cracked to give olefins and smaller paraffins.
- Olefins are cracked to give smaller olefins
- Naphthenes are cracked to give smaller olefins.
- Alkyl aromatics lose their side chains as olefins, leaving benzene.

Important Secondary Reactions are :

- Isomerization of olefins to iso-olefins
- Hydrogen transfer between naphthenes and olefins, forming aromatics and paraffins.

The acid catalyzed cracking reactions form the main reactions pathway for light gas, LPG components and long chain paraffin components of diesel. Isomerization reactions give important high octane components in the gasoline. Hydrogen transfer reactions improves gasoline yield and stability. These reactions produce paraffins and aromatics that have low octane ratings. Dehydrogenation reaction occurs due to presence of metals, such as nickel, vanadium, on the catalyst. Dehydrogenation and dealkylation due to metal sites on catalyst, produced large amount of H₂ and paraffin components with low octane ratings (Kiran Pashkanti et al 2011)[8]. In the cracking process carbon is also produced which gets deposited on the catalyst as catalyst coke.

The modeling of cracking kinetics is based on lumping of compounds. The lumps are made on the basis of boiling range of feed stocks and corresponding products in the reaction system. The other approach is that in which the lumps are made on the basis of molecular structure. The number of kinetic constant depend on the number of lumps and these kinetic parameters are estimated considering the conversion of one lump to the other.

3.2 Riser Hydrodynamics

The riser reactor of a FCC unit can be divided into equal size compartments along the axis. The entry part of riser reactor consists of three phases catalyst(solid phase),hydrocarbons vapors and atomizing steam(gas phase),and hydrocarbons liquid .In the feed injection zone at the bottom of the riser , the catalyst particles and liquid drops are accelerated upwards, the gas velocity also increases because of feed vaporization , lowers the density of the flowing system due to formation of lower molecular weight products on cracking of gas oil [23]. In the feed injection zone the hydrocarbon feed dispersed in the form of droplets by the feed nozzles system comes in contact with the hot regenerated catalyst. The liquid drops vaporize due to intimate contact between liquid drops and hot catalyst . The flow of feed oil, atomizing steam and hot generated catalyst results in velocity, temperature and concentration gradients of high magnitude in this zone [11].

Variation of gas velocity influence the axial and radial profile of catalyst volume fraction. Catalyst temperature decreases due to heat requirement for increasing sensible heat of feed, vaporization and cracking reactions. Catalyst activity also decreases due to deposition of coke on catalyst surface. In the entry zone the rate of vaporization of feed affects the performance of riser reactor. Feed in the liquid phase can not crack easily. Slow vaporization of feed require very high catalyst to vaporized feed ratio along with high catalytic activity and temperature in the entry zone. This factor may cause undesirable secondary cracking reactions. Effective feed atomization into fine drops leads to faster vaporization.

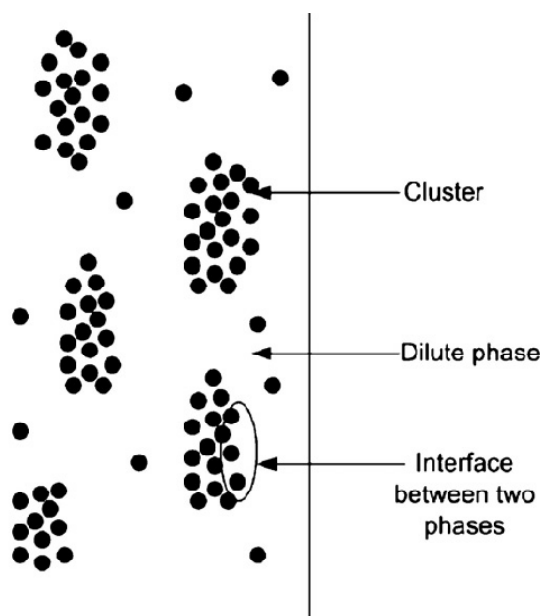


Fig. 8 .Schematic of the riser flow.

The feed injection design is important to control the flow of hydrocarbons at plug flow conditions for minimizing the temperature gradients in the inlet zone that causes undesirable cracking reactions. Increased number of nozzles at the bottom promotes proper contact of the feed oil with the catalyst particles. Feed nozzle design is also important which affect the performance of FCCU.

In the middle and upper section ,hydrocarbons vapors and solid catalyst are present since all the feed droplets vaporizes after travelling 2-4 m up from the feed inlet. In this section vapor stream carries with it the catalyst particles in suspension and there is some back mixing of these particles because of slip velocity between the solid and vapor phase. Catalyst particles are assumed to move as a clusters. Vapor density can be calculated using ideal gas law. The heat transfer is assumed to be mainly by convection mechanism in various proposed models. Heat transfer coefficients between gas and particles, solid to gas and gas to liquid are expressed as Nusselt number as a function of Reynolds number .Mass transfer of gas oil from gas phase to cluster phase and the products from cluster phase to gas phase occur due to concentration gradients between gas phase and cluster phase. Mass transfer coefficients can be expressed as the Sherwood number as a function of Reynolds number based on single particle diameter[23].

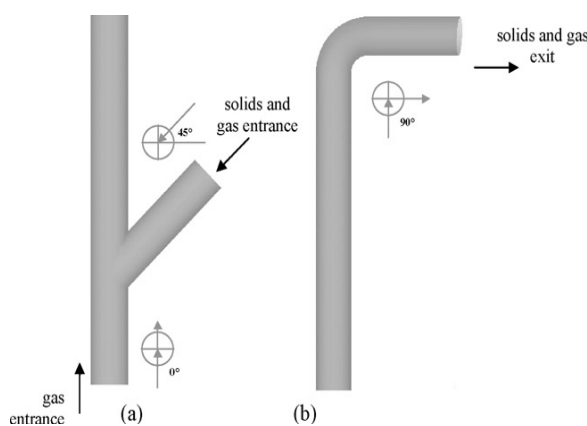


Fig.9. Entrance (a) and Exit (b) of a riser

The particles in a fluidized bed are so small that intra particles concentration and temperature gradients are negligible. Heat and mass transfer rates between fluid and particles are very high so that external temperature and concentration differences are negligible. The global rate is evaluated at bulk values of the temperature and concentration.

Core –annulus structure exists in FCC risers. At the bottom, generally there is higher concentration of solids and flow is highly nonhomogeneous. These in homogeneities tend to decrease as the gas phase accelerates and flow becomes more diluted. (Lopes,G.C et al,2011)[24]

The variation in velocity induces a higher concentration of gasoil at the centre of the riser whereas the catalyst concentration is high near the wall and this cause maximum conversion near the wall.

Ajay Gupta et al (2001 & 2003)[23,25] proposed a FCC riser model taking into account of the effect of feed atomization on conversion and yield. The developed model capable of predicting profiles of overall conversion, product yields, temperature, axial solid hold up , catalyst activity and other parameters. Result predicted that for a bigger drop size, catalyst circulation has to be decreased to achieve base case level coke yield. Lower catalyst circulation rate amount to lower heat withdrawal from the regenerator. This cause increase in regenerator bed temperature. On the riser side ,increase in drop size ,the conversion ,gasoline yield and gas yield exhibit a falling trend.

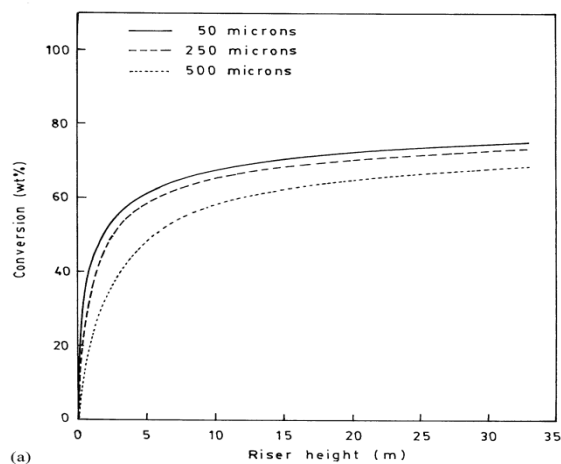


Fig. 10. Effect of Initial drop size on Conversion (Ajay Gupta et al 2001)

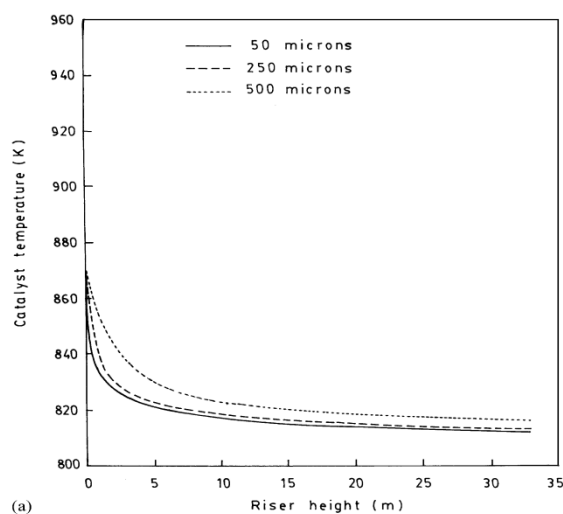


Fig. 11. Effect of initial drop size on catalyst temperature(Ajay Gupta et al,2001)

In the riser reactor higher overall conversion is predicted for smaller drop(Fig:10). Atomization of feed into fine drop facilitates high rates of heat transfer between catalyst and feed. Due to this temperature of catalyst fall in the entry zone as it losses heat to hydrocarbon feed. Large drops leads to high catalyst temperature in the riser due to low rate of heat transfer between catalyst and hydrocarbon feed (Fig.11).High temperature promotes secondary cracking of gasoline to coke. Higher cokes leads to fast deactivation, lower conversion and gasoline yield. (Ajay Gupta et al,2001)[25].

A.Gomez-Barea et al (2007)[26] studied mass transport effects during measurements of gas- solid reaction kinetics in a fluidized bed. In his study Dimensionless parameters, which govern the process , was derived from a simple two phase approach and an overall kinetic particle model. The inter phase effectiveness factor expressed in term of two observables one Weisz-Prater module and the product of Damkohler number at a particle scale and the other is particle effectiveness factor. The transport effects can be evaluated from these observables .The study primarily developed for the assessment of mass transport effects in isothermal systems.

But the methodology derived can be applied to other catalytic and non catalytic systems with some assumptions. D.Subbarao (2008)[27] proposed a model to explain gas to particle mass transfer in risers considering that particles move as clusters. The result indicate that gas to particle mass transfer coefficient decreases with increase in solid circulation flux for all the gas velocities and particle sizes. Mass transfer coefficients increases with increase in superficial velocity for the same solid circulation flux.

S.Das et al (2007)[28] in his study on FCC riser hydrodynamics found that with the increase in riser height and gas velocity, the voidage increases but at the end of the riser length, densification of solids occurs and voidage tends to decrease. An opposite trend a decrease in voidage all along the riser height, was noticeable with the rise in solids circulation rate.

S.I.Reshetnikov et al (1995)[29] developed a mathematical model for fluidized bed reactor considering unsteady state of the catalyst. In fluidized bed reactor, catalyst particles move continuously along a reactor height in the reaction medium with varying composition of reagents. Under the influence of the reaction mixture each particle permanently changes its properties being in unsteady state with respect to the surroundings reagents. The results indicate the effect on the temperature profile, conversion and efficiency of reactions.

Wang Shuyan et al (2008)[30] proposed a mathematical model for predicting gas and gasoline distributions in the cluster inside riser reactor. The results show that the molar concentrations of gas and gasoline of individual particle in the cluster were lower than those of an isolated particle in the stream due to low gas velocity through the cluster and low reaction rate inside the cluster. The reaction rates from VGO to gasoline, gas and coke of particles in the cluster were lower than that of an isolated particles. The reaction rates from gasoline to gas and coke of particles in the cluster were higher than that of the isolated particles.

Gabriela C. Lopes et al (2012)[31], presented a three dimensional, three phase flow model and a four lump kinetic base model to study the influence of different riser outlet bend designs on the dynamics of the flow and reactor efficiency. Results show that small changes in the outlet configuration had a significant on the flow patterns and on reactions yields. The use of abrupt exits results in solids, back mixing which enhanced the residence time of the catalyst and increased its concentration near the top of the riser. An area of densification is then created near the exit, which, depending on the severity of the exit restriction extends along the length of risers. The increase in the amount of catalyst particles favors the heterogeneous FCC reactions, which are proportional to the solids concentrations in the system. Due to this, risers with abrupt exits, had higher gasoline yield. The T-shape exit with a projected height of 0.4 m gave the highest value for this variable, indicate that this outlet configuration improves the reactor efficiency under applied conditions.

Ariane C.Barbosa et al (2013)[32] proposed a three dimensional CFD modeling of fluid catalytic cracking industrial riser with different geometric configurations to feed catalyst particles. The main objective was to study the effect of gas-particle turbulent mixing on thermal reaction and its effect on gasoline quality and yield in an industrial riser. In the proposed model, 10 lumps kinetic model of Jacob et al was modified to 11 lumps kinetic model by the addition of new lump named dry gas representing the product of the thermal cracking reactions. The results show that thermal cracking effects can be reduced by using an efficient mixture between the gas and solid phase inside the reactor.

Subramanya V. Nayak (2005)[33] proposed a model for simulating evaporation of a drop considering heat transfer from the gas phase as well from the solid particles. The simulated results indicate that influence of droplet diameter on riser performance is sensitive to the values of oil properties, used kinetics and operating conditions.

E.Leon Becerril et al (2004)[34] in his work studied the impact of modeling pressure gradient in the simulation of mono dimensional risers. The addition of a simplified momentum balance to the model, improved prediction of feedstock conversion and yield to products.

J.C.S.C Bastos et al (2008)[35] used a two-phase, 3-D CFD model to study radial solids velocity profiles in the riser of high flux circulating fluidized bed (HFCFB). The κ - ϵ two phase turbulence model was used to describe the gas solids flow in HFCFB. The model predicted a core-annulus flow in the dilute and developed flow regions except in the region of highest solids concentration at the level close to inlet.

Several other hydrodynamics CFD models have also been proposed that explain gas-solids flow and cracking reaction in FCC riser. Simulation were carried out to understand the influence of the various operating conditions on the performance of FCC riser reactor.

IV. Modeling of Regenerator

Regenerator modeling includes three aspects, hydrodynamics, combustion reaction kinetics, and temperature control. The regenerator is divided into two regimes namely an upper “dilute region” and a lower “dense region”. Dense region of the regenerator is further divided into a bubble phase and an emulsion phase (two phase model). The temperature in the emulsion phase is assumed to be higher than in bubble phase due to high density of catalyst in emulsion phase. FCC catalyst regeneration is complex process and many factors (Solid-mixing, inlet air velocity, temperature of spent catalyst, carbon content on spent catalyst etc.) influence the performance of the regenerator. (K.K.Dagde et al, 2013)[9]

4.1 Regenerator Kinetics

In the regenerator, the catalyst are regenerated by burning off the coke that deposit on catalyst during cracking reactions. Catalytic cracking yields large amounts of coke (4-8 wt % of feed). The activity of the catalyst also reduces by deposition of coke. Hydrogen rich compounds either volatilize at the temperatures of a typical regenerators (650-800 ° C) or crack to combustibles and coke.

During catalyst regeneration in FCC unit coke is burnt to produce carbon monoxide and carbon dioxide.

The following irreversible coke combustion occur in the emulsion-phase of a regenerator.



Where K_c is the reaction rate constant for coke burning, K_{CO} is the reaction rate constant for catalytic CO combustion. Last equation is called “after burning” reaction which takes place in the dense region if sufficient O_2 is supplied. The promoter usually a metal like Platinum is attached to FCC catalyst to initiate CO burning to CO_2 . By providing better dispersion of Platinum on catalyst, more effective combustion of coke can be achieved.

Minimum air velocity required to initiate combustion reactions in the regenerator for proper catalyst regeneration and complete burning of carbon monoxide to carbon dioxide, the inlet air velocity must be in excess of the “theoretical” or exact amount required.

In-Su-Han et al (2004)[10] studied fluidized catalytic cracking process under full and partial combustion mode. The results show that in the full combustion mode, an excess of combustion air is supplied to keep the coke & carbon monoxide at relatively low concentration in the regenerator. In partial combustion mode, the incomplete combustion of the coke prevails and thus results in high concentration of carbon monoxide and coke in the regenerator. The full combustion mode is preferred as conventional operating mode of FCC processes because the unit operated in this mode is typically more stable and controllable than in the partial combustion mode. The study results also indicate that operation in full combustion mode requires less operating costs than in the partial combustion mode therefore increase the net profits.

4.2 Regenerator Hydrodynamics

The spent catalyst from the separator enters into the regenerator unit where it is fluidized by the hot air blowing from the bottom of the regenerator. For modeling purpose different model were developed by the researchers like by dividing the fluidized bed into two beds model of different densities, the dense bed and dilute bed, grid effect model, two region model and bubbling model. H.A.Affum et al.(2011)[36] studied modeling and simulation of non ideal flow in a Fluid Catalytic Cracking (FCC) regenerator assuming mixed and plug flow regimes. The result show that overall coke on catalyst conversion of 88.85 % was obtained in the absence of dead volume in a mixed regenerator. The simulation result also shows that conversion of coke on catalyst is less efficient in plug flow reactor as compared to mixed flow reactor.

K.K.Dagde et. Al(2013)[37] developed a model which incorporate coke combustion kinetic and hydrodynamics. The results indicate that catalyst bed height and inlet air velocity have significant effects on catalyst regeneration. With increases in inlet air velocity the quantity of coke burnt increases, due to increase in the rate of oxygen supply to the regenerator. With increase of catalyst bed height, the concentration of oxygen in the regenerator increases partly due to incomplete combustion of coke and partly from its constant supply from inlet air, the concentration of carbon dioxide decreases, while the concentration of carbon monoxide increases. Temperature also play a important role in the regeneration of catalyst. The regenerator should be operated at an optimum temperature that gives high reaction rate of coke combustion and low carbon content on regenerated catalyst.

V. CONCLUSION

In this research paper various aspect related to Fluidized catalytic cracking has been explored through various review papers. Based on these papers, it has been concluded that kinetic modeling of the riser reactor is based either on the lumping scheme or on the 'single events approach. The kinetic constants evaluated using the lumping scheme are empirical in nature and are too much feed and plant specific. Due to instantaneous cracking reactions most of the cracking takes place in the first few meters of the riser height. Therefore, the riser hydrodynamics are complex in nature in the inlet zone and it needs detailed investigations. The yield of valuable product can be increased by using properly designed nozzles for feed atomization. Also due to the effect of feed atomization (drop size) on the overall conversion, gasoline yield and gas yield increases with decreasing drop size on riser side. FCC riser reactor model are dependent on the values of cracking reactions rate constant, which can easily be obtained with the help of kinetic model for different characteristics of the feedstock, type of catalyst, activity of catalyst and operating parameters. In FCC riser the gas and solids feed flow rates influence the flow pattern. A different interaction between the phases changes the way in which heat is transferred and consequently the yield of the reactions. Thus, it is important to know the non uniformities of the flow in order to correctly predict the reaction yield. It is necessary to maintain a more uniform radial dispersion to get a high yield of gasoline. The complete modeling of the hydrodynamics of system improves the prediction of the behavior of FCC risers. Heavy gas oil is adequately converted into gasoline and gases. Overall conversion achieved with the heavy gas oil is greater than that with light gas oil. Gases and coke yields are greater for Light gas oil. Zeolite and different modified form of zeolites are mainly used as FCC catalyst for improvement in gasoline yield, octane number enhancement, production of cleaner fuels and olefins productions. Minimum air velocity is required to initiate combustion reactions and proper catalyst regeneration in the regenerator.

ACKNOWLEDGMENT

The author gratefully acknowledge I K Gujral Punjab Technical University, Kapurthala (India) for providing guidance, Support and all other required facilities.

REFERENCES

- [1] A. Blasetti, H. De Lasa, "Heat Transfer Prediction in the Riser of a Novel Fluidized Catalytic Cracking Unit," *Ind. Eng. Chem. Res.*, vol 40(21), pp.4623-4632, 2001.
- [2] Bolas G.M., A.A. Lappas, D.K. Iatridis, I.A. Vasalos, "Five Lump Kinetic Model with selective catalyst deactivation for the prediction of the product selectivity in the fluid catalytic cracking process," *Catalyst today*, Vol 127, pp 31-43, 2007.
- [3] Elamurugan P., Dinesh Kumar D., "Modeling and Control of Fluid catalytic Cracking Unit in Petroleum Refinery", *International Journal of Computer Communication and Information System*, 2 No:1, ISSN:0976-1349, pp 55-59, 2010.
- [4] In-Su Han, Chang-Bock Chung, James B Riggs, "Modeling of a Fluidized Catalytic Cracking Process," *Computer and Chemical Engineering*, Vol 24, pp 1681-1687, 2000.
- [5] Bolas G. M., I. A. Vasalos, A. A. Lappas, D. K. Iatridis, G. K. Tsoni, "Bulk Molecular Characterization Approach for the Simulation of FCC Feedstocks", *Ind. Eng. Chem. Res.*, vol 43(13), pp 3270-3281, 2004.
- [6] Babita Behera, Siddarth S. Ray, I.D. Singh, "Structural Characterization of FCC feeds from Indian refineries by NMR Spectroscopy", *Fuel*, vol 87, pp 2322-2333, 2008.
- [7] J. Ancheyta - Juarez, S. Rodriguez Salomon, M.A. Valenzuela - Zapata, "Experimental Evaluation of Vacuum Gas oil-Light Cycle oil blends as FCC feedstock", *Energy and Fuel*, Vol 15, pp 675-679, 2001.
- [8] Kiran Pashikanti and Y. A. Liu, "Predictive Modeling of Large Scale Integrated Refinery Reaction and fractionation systems from plant data. Part 2: Fluid Catalytic Cracking (FCC) process", *Energy Fuels*, vol 25, pp 5298-5319, 2011.
- [9] K.K. Dagde and Y.T. Puyate, "Modeling and Simulation of Industrial FCC Unit: Analysis Based on Five Lump Kinetic Scheme for Gas - Oil Cracking", *International Journal Of Engineering Research and Application*, vol 2(5), pp 698-714, 2012.
- [10] Han In-Su, Riggs B. James, Chung Chang-Bock, "Modeling and Optimization of a fluidized catalytic cracking process under full and partial combustion modes", *Chemical Engineering and Processing*, vol 43, pp 1063-1084, 2004.
- [11] Raj Kumar Gupta, Vineet Kumar and V.K. Srivastava, "Modeling and Simulation of Fluid catalytic cracking unit: A review", *Reviews in Chemical Engineering*, Vol 21, No(2), 95-131, 2005.
- [12] Dewachtere N.V., Froment G.F., I. Vasalos, Markatos. N., Skandalis N., "Advanced Modeling of Riser Type Catalytic Cracking Reactors", *Applied Thermal Engineering*, Vol 17(8-10), pp 837-844, 1997.
- [13] Gupta Kumar Raj, Kumar Vineet, Srivastava V.K. (2007), "A New generic approach for the Modeling of fluid catalytic cracking (FCC) Riser Reactor", *Chemical Engineering Science*, Vol 62, pp 4510-4528, 2007.
- [14] Mehran Heydari, Habib Aleebhrabim and bahram Dabir, "Study of Seven Lump Kinetic model in the Fluid catalytic cracking Unit", *American Journal of Applied Sciences*, vol 7(1), pp 71-76, 2010.
- [15] Prabha K. dasila, Indranil Choudhury, Deoki Saraf, sawaran Chopra, ajay Dalai, "Parametric Sensitivity studies in a Commercial FCC Unit", *Advances in Chemical Engineering and Science*, Vol 2, pp 136-149, 2012.
- [16] Babatope Olufemi, Kayode Latinwo, Ayokunle Olukayode, "Riser Reactor Simulation in a Fluid Catalytic Cracking Unit", *Chemical & Process Engineering Research*, 7, 12-23, 2013.
- [17] H.A. Affum, I.I. Mumuni, G.K. Appiah, S.Y. Adzaklo, M.A. Addo, "On the conversion of gas oil in Fluid Catalytic Cracking Risers: Application of Residence time distribution (RTD) concept", *International Journal of Scientific & Technology Research*, Vol 2(10), pp 168-173, 2013.
- [18] R. Zafari, A. Nemati Kharat, "Application of Zeolitic Additives in the fluid Catalytic Cracking (FCC)", *Journal of Nanostructures*, vol 3, pp 209-217, 2013.
- [19] Thomas F. Degan, "Application of zeolites in petroleum refining", *Catalysis*, vol 13, pp 349-356, 2000.
- [20] Ziyuan Liu, Sheng-Li Chen, Xiujun Ge, Peng Dong, Jinsen Gao, and Zhiming Xu (2010), "Measurement of Diffusion Coefficient of Heavy oil in Fluidized Catalytic Cracking (FCC) Catalysts", *Energy Fuels*, vol 24, pp 2825-2829, 2010.
- [21] Rafael Maya - Yescas, Elizabeth Leon - Becerril, and Daniel Salazar - Sotelo, "Translation Of MAT kinetic data to Model Industrial catalytic cracking Units", *Chemical Engineering Technology*, vol 27(7), pp 777-780, 2004.

- [22] Rajasimman M,Prabhakaran M,"Micro Simulation studies on coked fluid catalytic cracking catalyst", International Journal of ChemTech Research,vol 1(3),pp 544-548,2009.
- [23] Gupta Ajay, D .Subba Rao,"Effect of feed atomization on FCC Performance :simulation of entire unit",Chemical Engineering Science ,vol 58,pp 4567-4579,2003.
- [24] Lopes G.C, Rosa L.M, Mori M, Nunhez J.R, Martignoni W.P,"Three-dimensional modeling of fluid catalytic cracking industrial riser flow and reactions", Computer and Chemical Engineering, Vol 35,pp 2159-2168,2011.
- [25] Ajay Gupta,D.Subba Rao,"Model for the performance of a fluid catalytic cracking (FCC) riser reactor :Effect of feed atomization",Chemical Engineering Science,Vol 56,pp 4489-4503,2001.
- [26] A.Gomez-Barea, P.ollero, B.leckner,"Mass Transfer effects during measurements of gas –solid reaction kinetics in a fluidized bed",Chemical Engineering Science,vol 62,pp 1477-1493,2007.
- [27] D.Subbarao,"A cluster model for Mass transfer in Riser",Journal Of Engineering Science and Technology,Vol 3(2),pp 131-137,2008.
- [28] S.Das(Bose), R.K.Saha, P.Sen Gupta,"FCC riser hydrodynamics: Effects of some operating variables", Indian Journal of Chemical Technology,vol 14,pp 473-480,2007.
- [29] Reshetnikov S.I ,Balzhinimaev B.S,Gaevoi V.P,Ivanov A.A,"Mathematical modeling of a fluidized bed reactor taking into account unsteady state of the catalyst", The Chemical Engineering Journal ,60, 131-139,1995.
- [30] Wang Shuyan,LU Huilin,GAO Jinsen,XU Chunming and SUN Dan,"Numerical Prediction of Cracking Reaction of Particles clusters in Fluid Catalytic Cracking Riser reactors",Chinese Journal of Chemical Engineering,Vol 16(5),pp 670-678,2008.
- [31] Gabriela C.Lopes,Leonardo M.Rosa,Milton Mori,Jose R. Nunhez and Waldir P.Martigoni,"CFD study of Industrial FCC Risers :The effect of outlet configuration on hydrodynamics and Reactions",International Journal of Chemical Engineering,ID:193639,16 pages,DOI:10.1155/2012/193639,2012.
- [32] Ariane C.Barbosa,Gabriela C.Lopes,Leonardo M.Rosa,Milton Mori,Waldir P.Martignoni,"Three dimensional Simulation of catalytic cracking reactions in an Industrial scale riser using a 11 lump Kinetic Model,"Vol 11,AIDIC conference series,The Italian Association of Chemical Engineering, DOI:10.3303/ACOS 1311004,2013
- [33] Subramanya, V.Nayak, Saket L.Joshi,Vivek V.Ranade , "Modeling of Vaporization and cracking of liquid oil injected in a gas –solid riser",Chemical Engg. Science,vol 60,pp 6049-6066,2005.
- [34] E.Leon-Becerril,R.maya-Yescas ,D.salazar-Sotelo,"Effect of modeling pressure gradient in the simulation of industrial FCC risers",Chemical Engineering Journal ,vol 100,pp 181-186,2004.
- [35] J.C.S.C.Bastos,L.M.Rose,M.Mori,F.Marini,W.P.Martigoni,"Modeling and simulation of a gas-solids dispersion flow in a high flux circulating fluidized bed(HFCFB)",Catalysis Today,vol 130,pp 462-470,2008.
- [36] H.A.Affum, P.S.Adu, C.P.K.Dagadu, M.A.addo, I.I.Mumuni, G.K.Appiah, A.Coleman and S.Y.Adzaklo,"Modeling Conversion in a Fluid Catalytic Cracking Regenerator In Petroleum Refining",Research Journal of Applied Sciences, Engineering and technology,Vol 3(6),pp 533-539,2011.
- [37] K.K.Dagde and Y.T.Puyate , "Modelling Catalyst regeneration in an Industrial FCC unit",American Journal Of Scientific and Industrial Research,Vol 4(3),pp 294-305,2013.