MATHEMATICAL MODELING OF FCC RISER REACTOR AND COMPARISON WITH DOWNER REACTORS

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Abstract: The aim of this study is to develop a one dimensional hetrogeneous two phase model for the adiabatic FCC reactor assuming a plug flow for both the solid and gaseous phases. The reaction kinetics of the cracking reactions are represented by the four lump kinetic scheme. The hydrodynamics of the riser are considered taking the slip factor into account. The model consists of ordinary differential equations is solved in order to get the yield patterns and study the effect of process variables like input catalyst temperature, and catalyst oil ratio on the yield of gasoline. Another model is developed for the downer reactor using hydrodynamic modeling, mass and energy balances. The model developed is solved using Runge- kutta IV order implemented in MATLAB and the differences in the axial profiles and gasoline yield as compared with the riser reactor are discussed.

Key words: FCC, four lump kinetic scheme, slip factor

INTRODUCTION

The world today is observing a general increase in demand of desirable light refinery products. But due to decrease in quality of crude oil and damage caused by inferior products to the environment there is a need to invest in research in order to find new hydro processing units, which can be integrated with older refineries also. FCC is a process in which the heavier feed stock like gas oil is cracked into the lighter and more desirable products, thus helping in meeting the market demands. This calls for more optimized and controlled processes and hence use of advanced process engineering tools for process planning has become necessary. A FCC unit typically comprises of a reactor, a regenerator and fractionating column to separate the products. The reactor can be of two types: riser and downer. The riser reactors have been used widely in the past years and many authors have tried to develop a model for the same (Ali and Rohani 1997, Arbel et al. 1995, Han and Chung 2001 and Theologos and Markatos 1993) while the very few attempts have been made to model the downer reactors. The main difference between the two kinds of reactors lies in the simple fact that in a riser reactor the feed and the catalyst particles are injected from the bottom of the reactor using dispersing steam which is used to fluidize the catalyst particles while in a downer reactor the feed and catalyst particles enter from the top and flow in the direction of gravity. The difference in the flow of the gas and solid phases inside the riser and downer reactors results in a variation in the yield and axial profiles of the products. As the feed comes in contact with the hot catalyst from the regenerator it is vaporized and cracked into lighter products like gasoline, light gases and coke. The coke thus produced gets deposited on the catalyst surface leading to catalyst deactivation and the catalyst particles are sent to the regenerator where the coke is burnt off from the catalyst surface. This regenerated catalyst is re injected into the reactor.

Modeling of FCC reactor broadly involves material balance, hydrodynamic modeling, kinetic modeling and catalyst deactivation model. Many empirical equations and catalyst deactivation model are available from the previous works (Pitault et al. 1995, Weekman 1979, Gianetto et al., Corella et al. 1985) which can be used to predict the deactivation of catalyst caused by coke deposition on catalyst surface. The kinetic modeling is done by grouping the chemical species into lumps based on the boiling point range. The three

lump kinetic model was first developed by Weekman et al. and considered gasoil, gasoline and light gases and coke as the three lumps. Since this model did not consider the coke as a separate lump a higher lump model was needed to predict the coke yield. Hence a four lump model was developed by Lee et al. in 1989 where gasoil, gasoline, light gases and coke were the four lumps considered. In literature kinetic models ranging from 3 to 19 lumps are available. The most widely used models are the 4 and 10 lump model.

Several studies have been carried out to form dynamic models for FCC riser reactor in order to increase the product yield and selectivity and enhance catalyst activity. The feedstock entering the bottom of the riser reactor is vaporized in the presence of regenerated catalyst and all the cracking reactions occur in vaporized state inside the riser reactor, as the catalyst particles and hydrocarbon droplets and vapor move upward along the riser length. Hence a lot of complex phenomenon take place simultaneously inside the riser reactor and need to be taken into account while modeling. Many authors have considered same velocities for both the phases in the riser reactor. But it is advisable to consider a slip between the two phases. For this a hydrodynamic model taking slip factor into account is used in the present paper (Patience et al. 1992).

FCC riser reactors have been widely used in the past for various industrial processes but another type of circulating fluidized bed reactors i.e. the downer reactors have been used in the recent years. An attempt to study and compare the yield patterns for both types of reactors has been made in the present paper. The development of downer reactor mainly comprises of mass and energy balance, kinetic model along with a hydrodynamic model for the two phases. The downer reactor provides a uniform contact time between the gas and solid phases.

MODEL ASSUMPTIONS

Since a number of complex phenomenon's occur simultaneously inside the FCC reactor a number of assumptions are made to simplify the model developed:

- One-dimensional transported plug flow reactor
- radial and axial dispersion inside the reactor are assumed to be negligible
- Constant heat capacities and viscosities are assumed for all components
- Negligible adsorption and dispersion effects are considered inside the catalyst particles
- The pressure changes that occur along the reactor length are assumed to be due to the static head of catalyst particles in the riser
- It is assumed that the coke on the catalyst does not affect flow of the fluid inside the riser reactor
- The catalyst and gas temperature is assumed to be same in each section of riser
- The catalyst and coke are assumed to have the same specific heat
- The dynamics of a reactor is assumed to be in quasi- steady state
- Instantaneous vaporization of feed

MATERIAL AND ENERGY BALANCE

The riser is divided into small cross sectional areas of infinitely small length. Each volume element of the riser is assumed to have two phases a) solid phase b) gaseous phase. Both the phases are assumed to be perfectly mixed with no heat and mass transfer resistances. For each section of the riser reactor mass balance for each individual chemical species, energy balance and hydrodynamic studies are carried out.

Taking the above mentioned assumptions into consideration following equations can be used to determine the mass balances in the FCC reactor:

$$F_z - F_{z+\Delta z} + (A*\Delta z)*\rho_{cat}*(1-\varepsilon)*r_{iz} = 0$$

The above equation gives material balance for a component i between position z and $z+\Delta z$.

$$\frac{dy_i}{dz} = \frac{A*(1-\varepsilon)*\rho_{cat}*r_i}{F_g}$$

$$F_g = F_{go} + F_{ds}$$

Energy balance:

This equation is used to calculate the mixing temperature at inlet of the riser:

@ Z=0

$$F_{cat}*Cp_{cat}*(T-T_{cat}) + F_{go}*Cp^{l}_{go}*(T_{vap}-T_{go}) + F_{go}*Cp^{v}_{go}*(T-T_{vap})$$
$$+F_{go}*\Delta H_{vap} + F_{ds}*Cp_{ds}*(T_{ds}-T) = 0$$

Where T is the inlet temperature

@ Z=h

$$(F_g*Cp_g + F_{cat}*Cp_{cat})*\frac{dT}{dz}$$
$$= \sum_{i=1}^{n} r_i*\Delta H_i*A*(1-\varepsilon)*\rho_{cat}$$

Pressure Balance

$$-\frac{dP}{dz} = \rho_{cat}*(1-\varepsilon)*g$$

The density of the gaseous phase is calculated using the ideal gas law:

$$\rho_g = \frac{P*MW_g}{R*T}$$

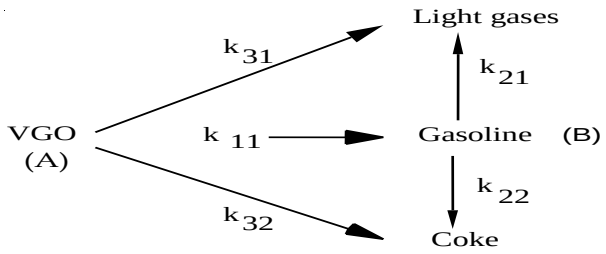
The average molecular weight of the gaseous phase is given by:

$$MW_g = \sum_{i=1}^{4} y_i*MW_i$$

KINETIC MODEL

A four lump kinetic model is used in this paper to represent the kinetics of cracking reactions.

Figure 1: Representation of a four-lumped scheme



This scheme considers the presence of four lumps namely: gasoil or the feed, gasoline, coke and light gases. The reaction orders for various cracking reactions which take place in the vapour phase are predicted.

a-) gas oil consumption rate: $-r_{go} = k_{11}\varphi C_{go}^2$

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

c-) light gases formation rate: $r_{lg} = k_{31}\varphi C_{go}^2 + k_{21}\varphi C_g$

d-) coke formation rate: $r_c = k_{32}\varphi C_{go}^2 + k_{22}\varphi C_g$

The kinetic parameters for the cracking reactions such as the reaction rate constants (K_i) and the catalyst deactivation function ϕ need to be estimated. All the kinetic parameters for the four-lump scheme cracking reactions are given in Table 1. Arrhenius equation is generally used to give the relation stating dependency of kinetic rate constants on temperature.

$$K_i^{'} = K_{io} * \exp(-E_i / RT)$$

where the values indicated by prime as predicted by eq. 11 is the kinetic constant of the reaction , K_{io} the pre exponential kinetic constants for the respective lumps, and E_i are the activation energies of the different lumps. The following equations give the expressions for rate constants (K_i) and the stoichiometric coefficients (V_{ij}) of the various lumps

$$K_i = K_i^{'} * V_{ij}$$

$$V_{ij} = M_i / M_j$$

As discussed earlier the catalyst deactivation occurs due to deposition of coke on the catalyst surface. In the present work the activity factor - ϕ which is a function of coke concentration on catalyst.

$$\phi = \frac{B+1}{B+\exp(A*C_{ci})}$$

The values for the parameters used in the above mentioned correlation for deactivation of catalyst; A and B as given in previous works is 4.29 and 10.4, respectively. (Pitault et. al., 1995).

HYDRODYNAMIC MODEL

For Riser Reactor:

To consider the slip factor arising due to difference in gas and solid flow, the correlation developed by Patience et al. is used in the present work. According to the correlation the gas interstitial velocity to average particle velocity ratio gives the value of slip factor and can be determined numerically. The empirical formula developed for the same is given as follows:

$$\Psi = \frac{U_o}{\varepsilon * U_p} = 1 + \frac{5.6}{Fr} + 0.47 Fr_t^{0.47}$$

Where $Fr = \frac{U_o}{\sqrt{g*D}}$ and $Fr_t = \frac{U_t}{\sqrt{g*D}}$

The gas superficial velocity is given by $U_o = \frac{F_g}{A*\rho_g}$

U_p, the average particle velocity in the riser used in eq. 6 can be calculated using eq. 7

$$U_p = \frac{G_s}{\rho_{cat}*(1-\varepsilon)}$$

Combining and solving eqs. 6 and 7 gives a new equation (eq.8) for calculating the average voidage in each section of the riser. It can be seen that average voidage is dependent upon solid mass flux, catalyst physical properties, superficial gas velocity and rise diameter which happen to be known quantities in FCC operation.

$$\varepsilon = 1 - \frac{G_s * \Psi}{U_o * \rho_{cat} + G_s * \Psi}$$

For downer reactor:

In order to study the hydrodynamics of the downer reactor, force balance is applied on a single particle of the gaseous phase stream. The equation developed is as follows:

$$\rho_p * V_p * \frac{\partial U_p}{\partial t} = \frac{1}{2} * \rho_g * \left| \frac{U_o}{\varepsilon} - U_p \right| * \left(\frac{U_o}{\varepsilon} - U_p \right) * A * C_D + (\rho_p - \rho_g) * V_p * g$$

In the equation mentioned above the inertial force is represented in terms of the drag force, gravitational force, and buoyancy force. The sign and value of slip velocity which can be given as the difference between the average particle velocity and interstitial gas velocity determines the direction in which the drag force on the particles act. Hence a modulus sign is used in the above equation. The drag coefficient, C_D, is the only empirical parameter used in this model.

$$C_D = \frac{18.5}{\mathrm{Re}_p^{0.6}}$$

$$\mathrm{Re}_p = \frac{\rho_g * \left| \frac{U_o}{\varepsilon} - U_p \right| * d_p}{\mu_g}$$

$$G_s = \rho_{cat} * (1 - \varepsilon) * U_p$$

If the, U_t, terminal velocity of any particle is known then the d_p, particle diameter can be calculated in terms of equivalent diameter for a spherical particle. In the downer reactor when the solids flowing in the downward direction have achieved a steady state i.e. the flow is fully developed then slip velocity is used to represent the terminal velocity for that particle. At a particular operating conditions the slip velocity can be determined as a function of gas superficial velocity, particle velocity and voidage given that the flow is completely developed.

$$U_s = \frac{U_o}{\varepsilon} - U_p$$

The slip velocity of a particle is equal to the terminal settling velocity of a particle falling through a fluid and is given by:

$$U_s = U_t = \sqrt{\frac{4 * d_p (\rho_p - \rho_g) * g}{3 * \rho_g * C_D}}$$

In the above equation for a given gas superficial velocity, voidage and average particle velocity are not known. The additional correlations required are provided by the mass balance equations which help in determining the unknown parameters.

DATA

The values of kinetic parameters, thermodynamic properties are obtained from Ahari et al., 2008 and are given in the following tables.

TABLE 1: Kinetic parameters

	K_0	E(kJ/mol)	ΔH (kJ/kg)
Gas oil to Gasoline	$1.15 * 10^3$	59.66	393
Gas oil to light gases	7.36*10	47.82	795
Gas oil to coke	1.79	30.95	1200
Gasoline to light gases	$4.26 * 10^2$	68.83	1150
Gasoline to coke	$5.99 * 10^{-4}$	57.74	151

TABLE 2: Industrial riser operating conditions used

	Case I	Case II	Case III	Case IV
Feed rate (kg/s)	19.95	25.7	26.9	23.6
Feed quality (API)	22.28	21.76	22.98	22.73
COR(kg/kg)	7.2	6.33	5.43	6.07
Inlet Pressure(kPa)	294	294	294	294
Feed Temperature(K)	494	494	494	494
Catalyst Inlet temperature(K)	960	1033	1004	1006
Steam (Wt %)	7	5.5	5	5.75
Steam Temperature(K)	773	773	773	773

TABLE 3: Riser Dimensions

	Height(m)	Diameter(m)
Riser/downer reactor	33	0.8

TABLE 4: Properties

Species	MW(kg/kmol)	C_p(kJ/kg.K)
Gas oil	333	2.67(Liquid), 3.3(Gas)
Gasoline	106.7	3.3
Light Gases	40	3.3
Coke	14.4	1.087
Steam	18	1.9
Catalyst	N/A	1.087

TABLE 5: Thermodynamic properties of the feed

Gas oil vaporization temperature	698 K
Viscosity of gas	$1.4 * 10^{-5}$ N.s/m^2
Gas oil enthalpy of vaporization	190 kJ/kg

MODEL SOLUTION

The model developed consists of a system of ordinary differential equations. The equations are not stiff in nature. Hence Matlab tool ODE 45 is suitable for solving the system of equations. A variable step Runge-Kutta Method is used by this tool in order to solve the system of ordinary differential equations numerically. Hence a MATLAB code has been developed for the purpose. Operating conditions in the four cases are used to validate the model while the operating conditions given in case I are used for analyzing the effects of various process variables on gasoline yield and to compare the riser and downer reactors.

SIMULATION RESULT AND DISCUSSION

In order to validate the model the results obtained by solving the model for four cases of operating conditions. The results obtained and the deviation in the results is reported in Table 6. As can be seen by comparing the results with the actual plant data, a good agreement between the plant data and the model prediction is observed.

TABLE 6: Comparison of simulation results with plant data

Case 1:

	Plant	Calculated	%deviation
Gasoline (% wt)	43.88	47.5	8.24
Coke yield (%wt)	5.63	5.53	-1.776
Outlet temperature (K)	795	797.5	0.314

Case 2:

	Plant	Calculated	%deviation
Gasoline (% wt)	46.90	49.47	5.5
Coke yield (%wt)	5.34	5.27	-1.25
Outlet temperature (K)	808	802.95	-0.625

Case 3:

	Plant	Calculated	%deviation
Gasoline (% wt)	42.79	45.806	7.05
Coke yield (%wt)	5.43	5.34	-1.57
Outlet temperature (K)	805	797.98	-0.872

Case 4:

	Plant	Calculated	%deviation
Gasoline (% wt)	41.78	44.91	7.5
Coke yield (%wt)	5.69	5.58	-1.85
Outlet temperature (K)	806	802.3	-0.458

Comparison of riser and downer reactors

The figures 1 and 2 compare the yield of gasoline for a downer reactor and riser reactor respectively. As can be seen from the figures the downer reactor gives better conversion of the feedstock and higher yield for gasoline. It is observed that the yield in first five meters of the riser reactor is more than the yield in first five meters of the downer reactor. But along the length of the reactor the final yield of the downer reactor becomes greater than that of the riser reactor. This is because after first few meters the catalyst is deactivated and the extent of cracking reactions in the riser reactor reduces while in the initial section of the downer the holdup of catalyst is less. Most of the conversion in the downer takes place throughout the

length of the downer reactor as due to forward mixing the feedstock comes in contact with a fresh catalyst at any axial location thus resulting in a higher yield at a later stage.

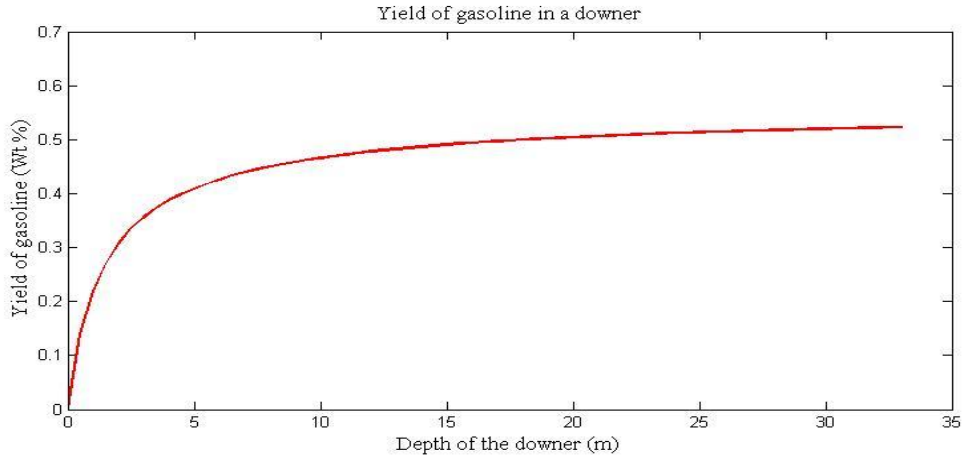


FIGURE 1: Yield of gasoline in a downer reactor at 960 K

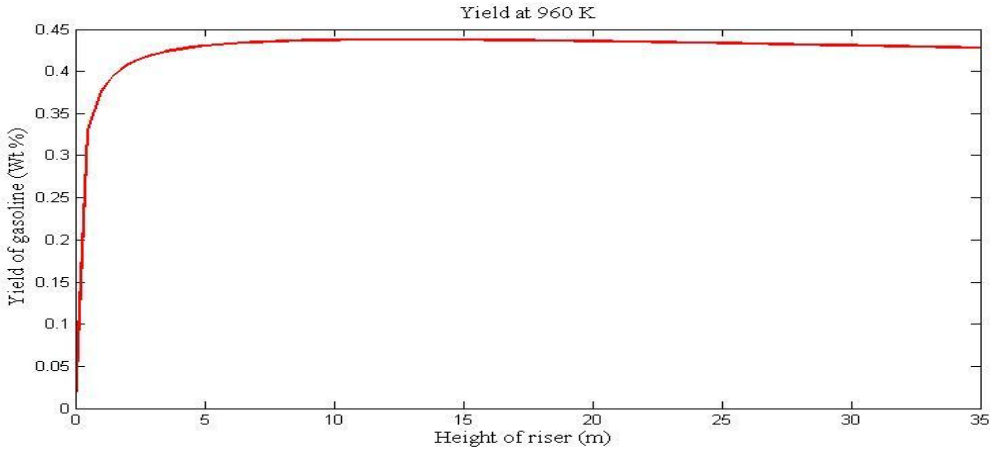


FIGURE 2: Yield of gasoline in a riser at 960 K

Figure 3 and 4 compare the yields of coke in a riser and a downer reactor respectively. As expected the downer reactor gives a lower coke yield as compared to the riser reactor, thus indicating a reduced level of secondary cracking reactions. In a downer reactor the catalyst particles reach a terminal velocity after some time which is higher than the gas phase velocity. Thus as the gas phase flows down it comes in contact with the fresh catalyst thus reducing the secondary cracking reactions and hence the amount of coke. On the other hand in a riser reactor the gas phase comes in contact with the deactivated catalyst after some time and a higher level of secondary reactions takes place.

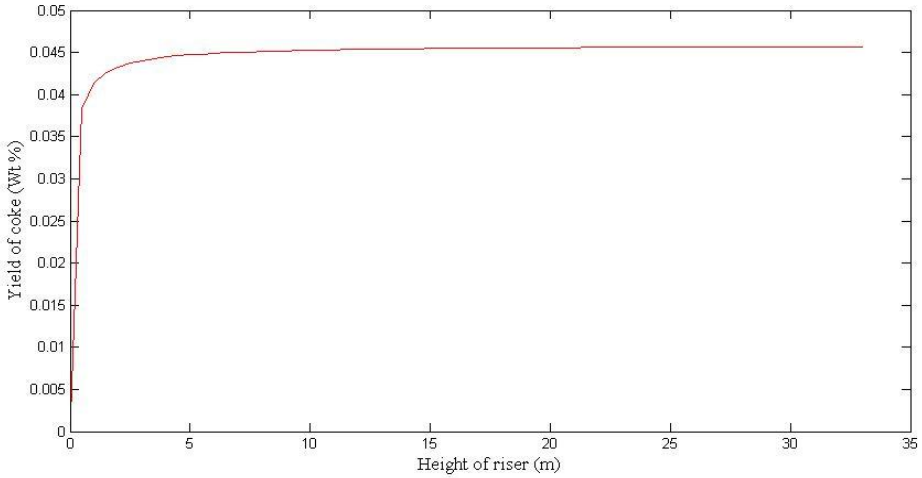


FIGURE 3: Yield of coke in riser reactor

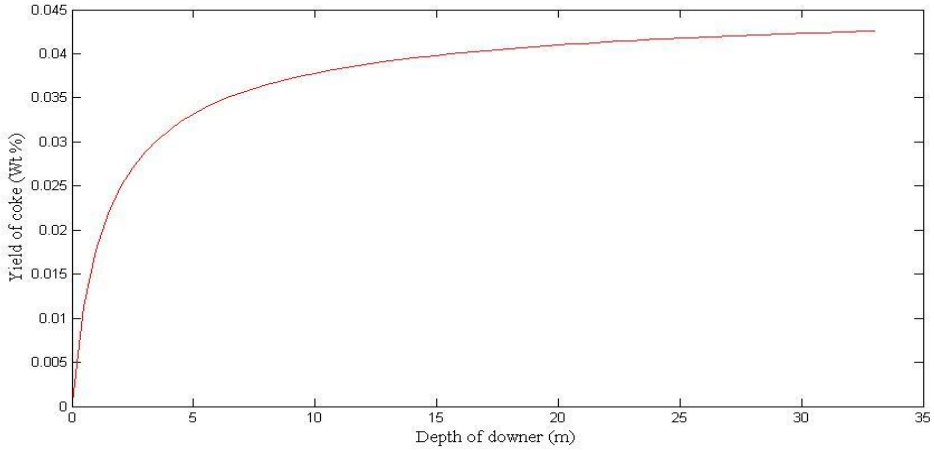


FIGURE 4: Yield of coke in downer reactor

Figure 5 gives the axial profiles of various chemical species present inside the riser reactor. According to the figure the maximum conversion of gasoil occurs in the initial section of the riser reactor i.e. the first five meters of the reactor. This fact is in accordance with what most of the authors like Ali et. al. 1997, Arbel et. al., Mckeen et. al have predicted till now. This can be reasoned based on the fact that:

- The concentration of regenerated catalyst at the bottom of the riser is very high. Also since the catalyst in bottom section of the riser is at a much higher temperature than in any other section of the riser. Hence due to high catalyst activity the rate of reaction is high.

- Also the concentration of feedstock i.e. gas oil vapor is maximum at the base of the riser as compared to any other section of the riser reactor where due to the reaction and molar expansion of gaseous phase the gas oil concentration decreases. Hence the rate of the reaction and therefore gasoil conversion is highest at the bottom of the riser.

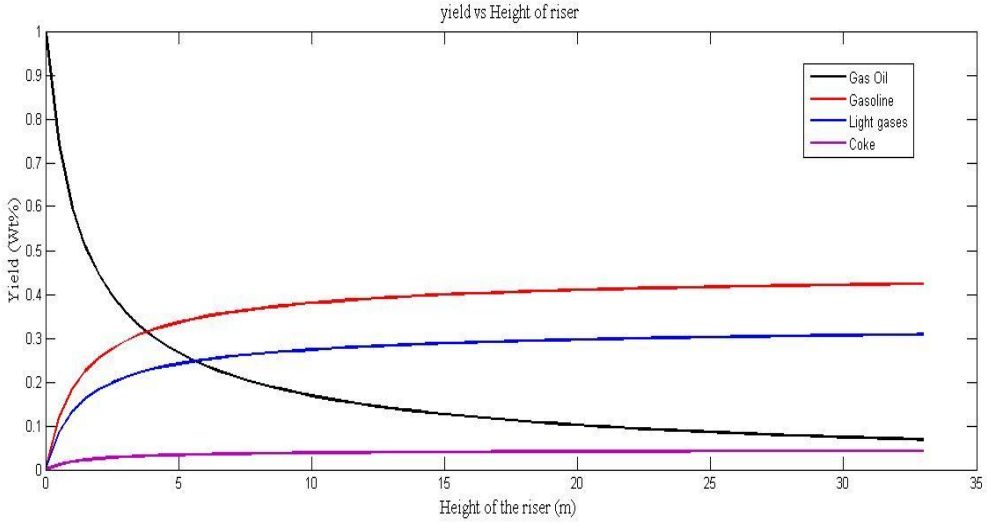


FIGURE 5: Yield patterns of Riser reactor

The following three figures give the temperature profile, the yield pattern of the species present and the pressure profile respectively. The temperature profile shown in figure 6 is decreasing in nature. This type of behavior can be reasoned by the endothermic nature of the reactions. Also it is observed that a rapid decrease in riser temperature takes place in first few meters of the riser length, thus accounting for the fact that most of the cracking reactions take place in first few meters of the riser length.

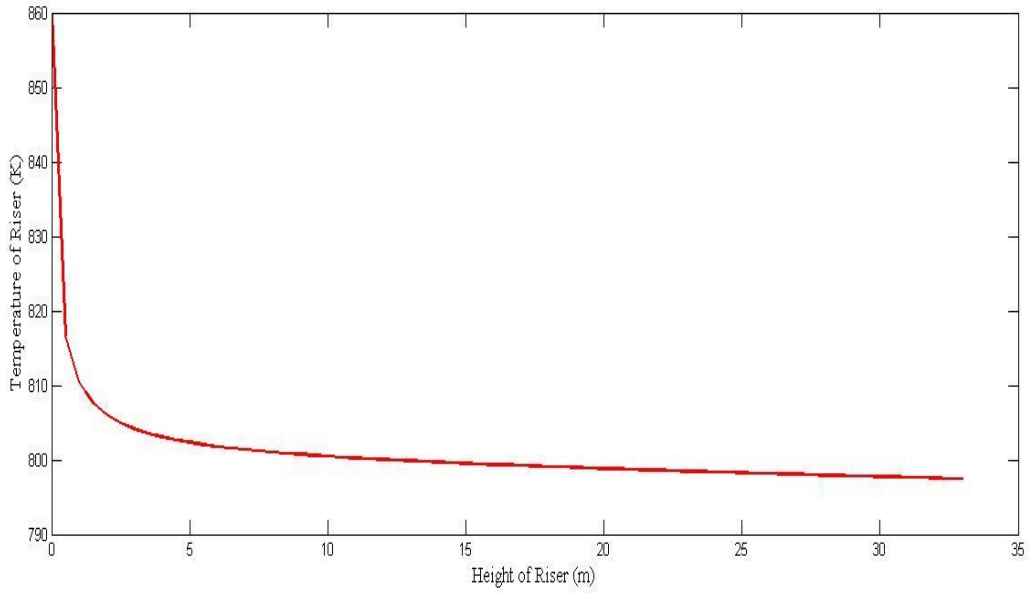


FIGURE 6: Temperature of riser v/s height of riser

Effect of Input catalyst temperature

Figure 7, 8 and 9 shows the axial profiles of gasoline at input catalyst temperature of 960K, 860K and 760K respectively. As can be seen from the three figures the yield of gasoline decreases with decrease in catalyst temperatures. Also a higher rate of reaction is observed for a higher catalyst temperature.

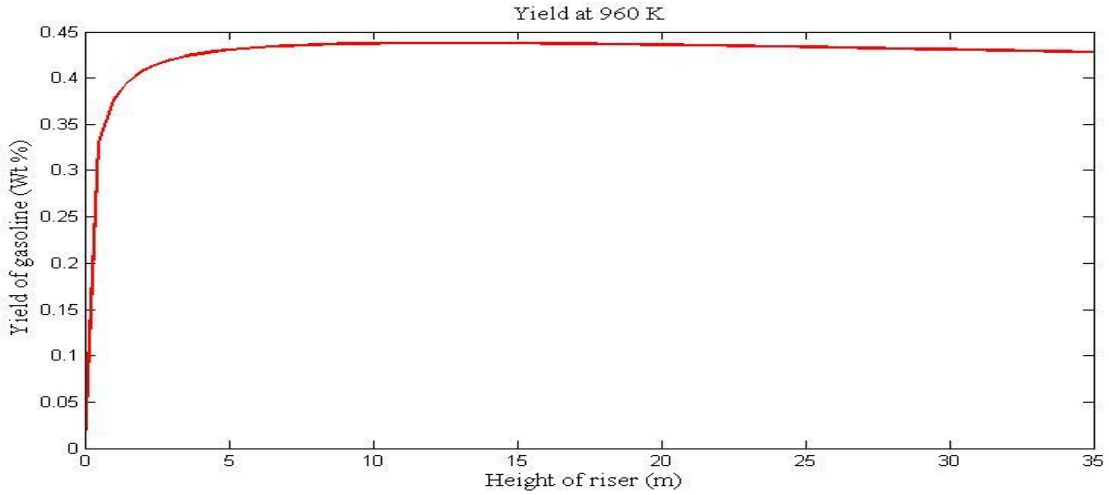


FIGURE 7: Gasoline yield at 960K

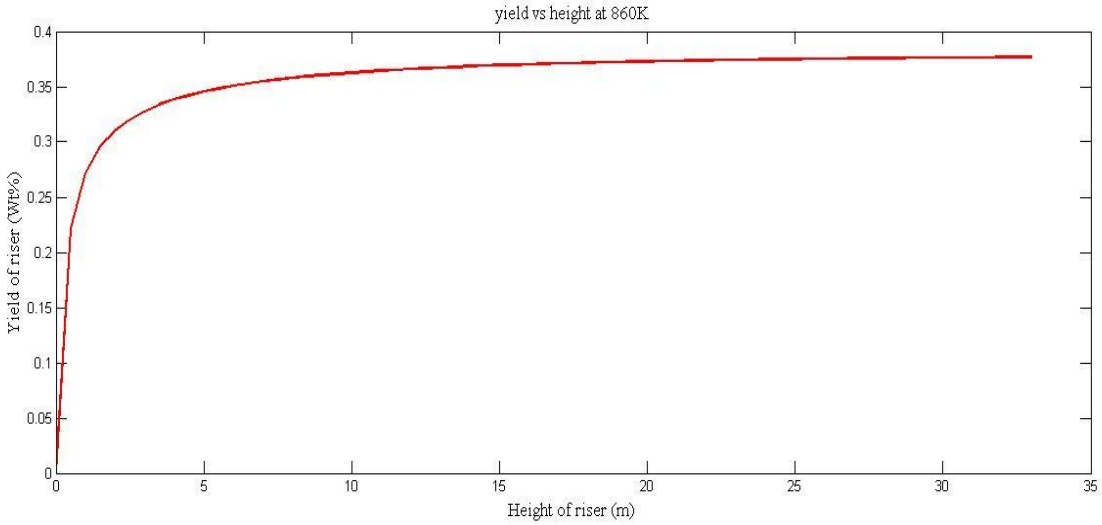


FIGURE 8: Gasoline yield at 860K

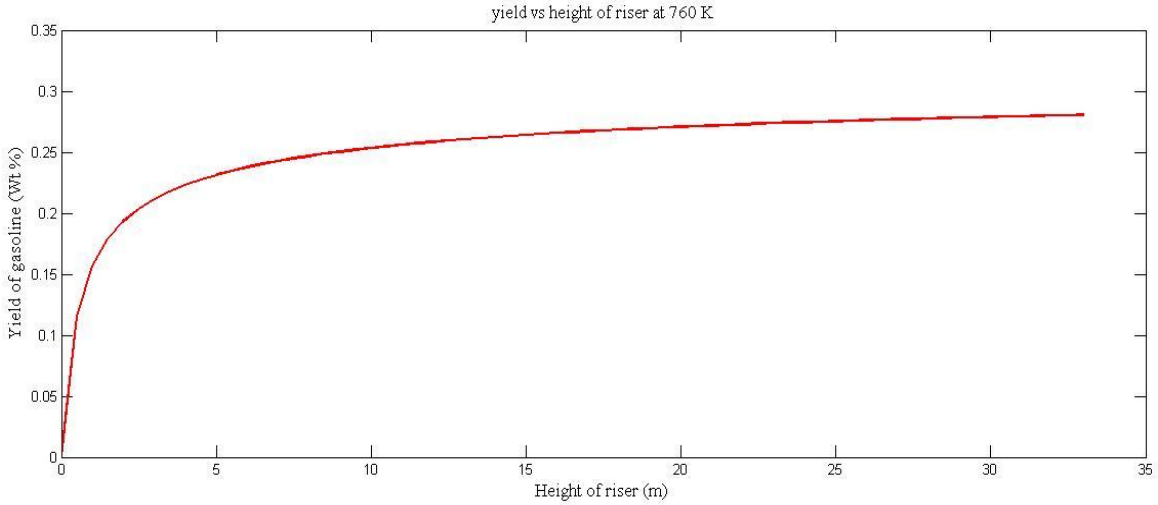


FIGURE 9: Gasoline yield at 760K

Effect of Catalyst to oil ratio (COR)

In order to study the effect of change of catalyst oil ratio on the yield of gasoline and this happens to be an important parameter for the FCC process. With increasing the COR at a constant catalyst temperature, the catalyst hold up (1-ε) increases with increasing of COR, leading to a higher conversion of the feedstock. Since due to cracking of the feedstock gasoil into lighter molecules leads to an increase in the interstitial velocity of the gas and catalyst particles, thus decreasing the residence time with further increase in value of COR. Figure 10, 11, 12 and 13 depict the yield of gasoline for different values of catalyst oil ratio ranging from 3, 5, 7 and 13 respectively at a input catalyst temperature of 960 K.As can be seen from the figures on increasing the value of COR from 3 to 5 the yield of gasoline increases and on further increasing the value of COR to 13 a decline in gasoline yield can be observed. This is because due to a higher rate of reaction due to an increase in COR the production of coke increases, thus deactivating the catalyst and hence resulting in a decrease in gasoline production at a later stage.

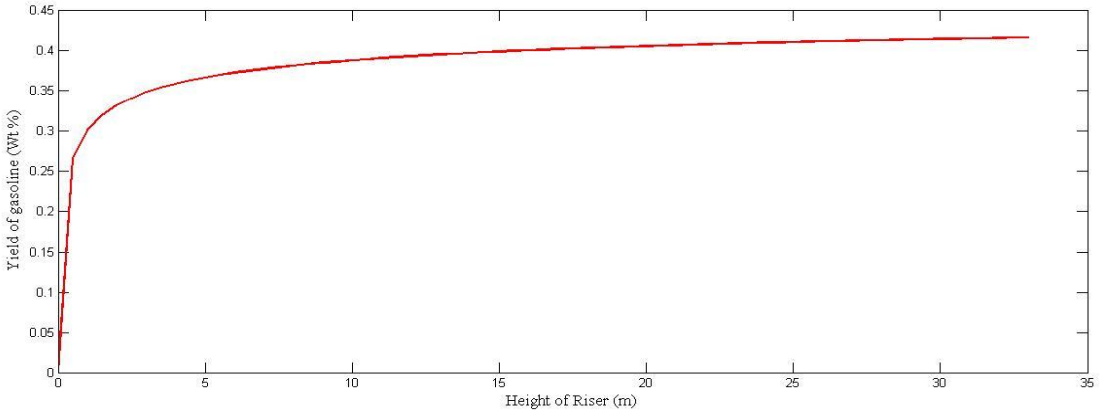


FIGURE 10: Yield of gasoline at COR=3

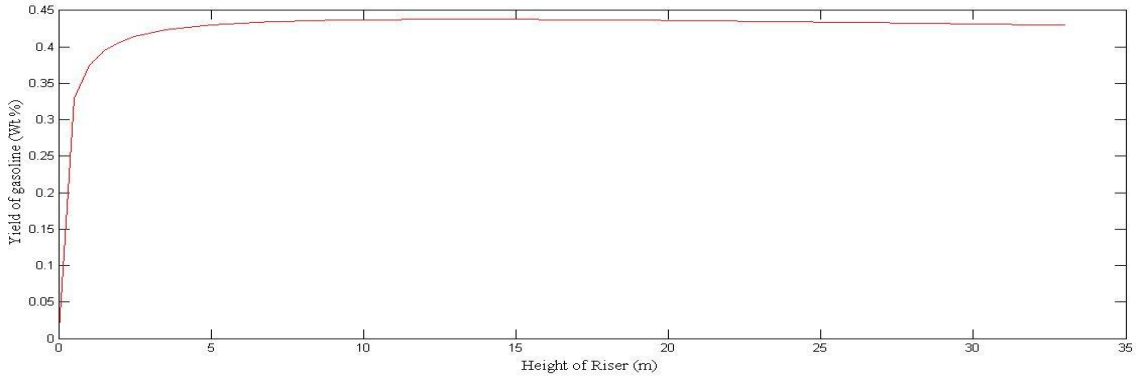


FIGURE 11: Yield of gasoline at COR=5

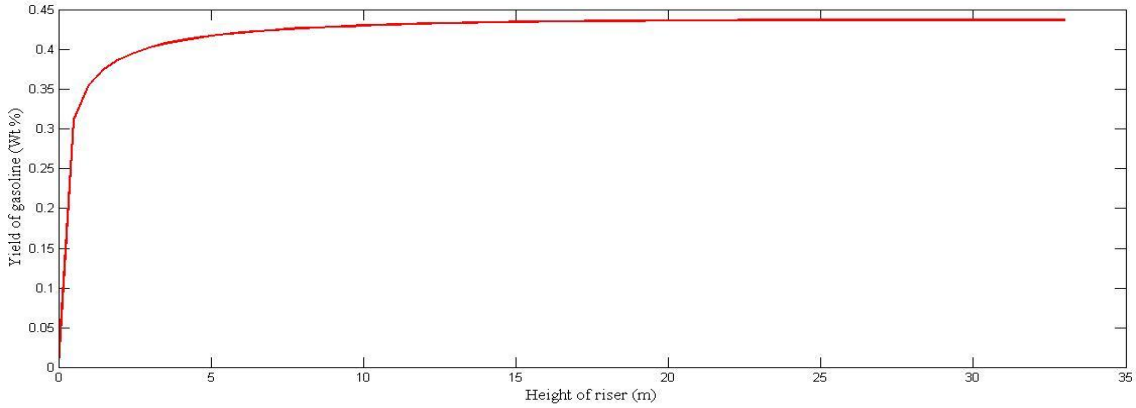


FIGURE 12: Yield of gasoline at COR=7

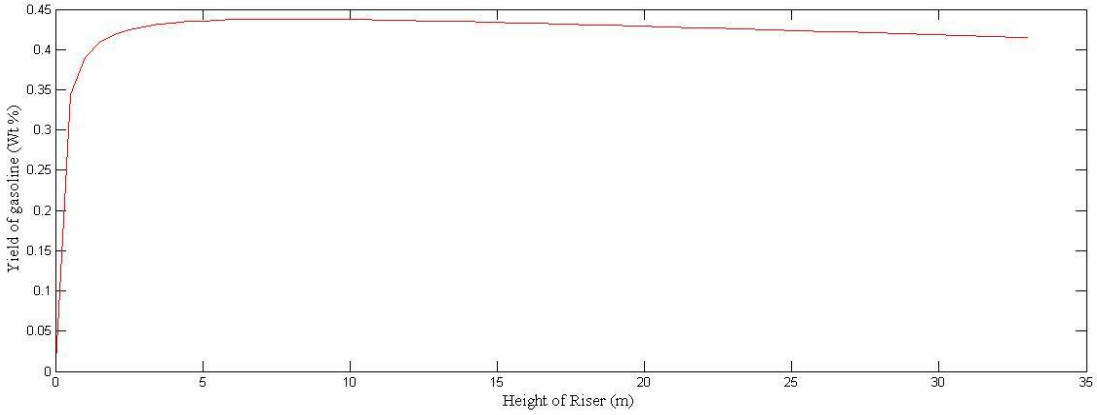


FIGURE 13: Yield of gasoline at COR=13

CONCLUSIONS

In this work a adiabatic one-dimensional model for FCC unit riser reactor was developed, that combines mass balance and riser hydrodynamic model for a four- lump kinetic scheme in order to predict the yield patterns. The yields predicted by the model were compared with four cases of plant data available and a good a agreement between the industrial data and simulation result is observed. The temperature, pressure and axial profiles of products are also presented. The effects of input catalyst temperature to riser, catalyst to oil ratio (COR) and feed rate on yield of gasoline, is analyzed. A comparative study of the yield patterns for a riser reactor and a downer reactor is also given. It was observed that the yield of gasoline increases with an increase in catalyst temperature as the rate of reaction is known to decrease with a decrease in catalyst temperature. The yield of gasoline was found to increase with increase in catalyst oil ratio till it reaches a maximum value and starts decreasing because after a certain level the production of coke is increased and hence due to deactivation of catalyst the gasoline yield starts decreasing. It can be seen that the yield of gasoline for a downer reactor is more than that obtained for a riser reactor even though the reaction rate in initial section of the riser reactor is more than that in the downer reactor. Also the yield of coke in downer is found to be less as compared to yield of coke in riser reactor due to reduced secondary reaction.

NOMENCLATURE

A: cross-sectional area of riser (m^2)
C_c: coke deposited on catalyst (wt%)
C_D: drag coefficient
Cp: specific heat (kJ/kg*K)
D: riser diameter (m)
d_p: particle diameter
E: activation energy (kJ/mole)
F: mass flow rate (kg/s)
Fr: Froude number
Fr_t: Froude number based on terminal velocity
g: gravitational constant 9.8(m/s^2)
G_s: solid mass flux ($kg/m^2.s$)
ΔH: heat of reaction(kJ/kg)
ΔH_{vap}: gas oil enthalpy of vaporization(kJ/kg)
K: Kinetic rate constant ($kg.s^{-1}.kgcat^{-1}$)
K_o: pre exponential factor 1st order ($kg,s^{-1}.kgcat^{-1}.wt^{-2}$);2nd order,($kg.s^{-1}.kgcat^{-1}.wt^{-1}$),
r: rate of reaction($kg.s^{-1}.kgcat^{-1}$)
R: gas constant 8.314 (J/mol.K)
Re: Reynolds number
MW: molecular weight (kg/kmole)
P: pressure (kPa)
T: temperature (K)
U_t: terminal velocity of the catalyst particle

U_o: riser superficial velocity (m/s)

U_p: average catalyst velocity (m/s)

U_s: slip velocity

V_p: average volume of a particle

y_i: mass fraction of component i (Wt %)

Z: axial position (m)

Greek symbols
α: coking constant
ε: void fraction
Ψ: slip factor
ρ: density (kg/m^3)
Φ: catalyst decay function
μ: viscosity

Subscripts
cat: catalyst
ck: coke
ds: dispersion steam entering the feed injection system
lg: light gases
g: gas
go: gas oil
gl :gasoline
p: particle
s: solid

Superscripts
v: vapor
l: liquid

REFRENCES

- Ahari Jafar Sadeghzadeh, Farshi Amir, Forsat Khaled, A mathematical modeling of the riser reactor in industrial FCC unit. Petroleum & Coal 50(2), 2008, 15-24
- Ali Hany and Rohani Sohrab, Dynamic Modeling and Simulation of a Riser-Type Fluid Catalytic Cracking Unit. Chem. Eng. Technol. 20, 1997, 118 – 130
- Arbel A., Huang Z., Rinard I. H., Shinnar R., Sapre A. V., Dynamic and Control of Fluidized Catalytic Crackers. 1. Modeling of the Current Generation of FCC's. Ind. Eng. Chem. Res.,34, 1995, 1228.
- Bolkan Yasemin, Berruti Franco, Zhu Jesse and Milne Bruce. "Hydrodynamic Modeling of CFB Risers and Downers". International journal of chemical reactor engineering, Volume I, 2003, Article A51
- Corella J.; Bilbao R.; Molina J. A.; Artigas A. Variation with Time of the Mechanism, Observable Order, and Activation Energy of Catalyst Deactivation by Coke in the FCC Process. Ind. Eng. Chem. Process Des. Dev. 24, 1985, 625.
- Fernandes J. L., Pinheiro C. I. C., Oliveira N. and Ribeiro F. Ramôa, Modeling and Simulation of an Operating Industrial Fluidized Catalytic Cracking (FCC) Riser. 2nd Mercosur Congress on Chemical Engineering, 4th Mercosur Congress on Process Systems Engineering, 2005
- Fligner M., Schipper P.H., Sapre A.V., Krambeck F. J., Two Phase Cluster Model in Riser Reactors: Impact of Radial Density Distributions on Yields, Chemical Engineering Science, Vol. 49, Nº. 24B,1994, 5813-5818
- Gianetto A., Farag H., Blasetti A., de Lasa H., Fluid Catalytic Cracking Catalyst for Reformulated Gasolines. Kinetic Modeling, Ind. Eng. Chem. Res. 33, 1994, 3053-3062.
- Gupta Ajay, Rao D. Subba, Model for the performance of a fluid catalytic cracking (FCC) riser reactor: effect of feed atomization. Chemical Engineering Science 56, 2001, 4489–4503
- Han In-Su, Chung Chang-Bock, Riggs James B., Modeling of a fluidized catalytic cracking process. Computers and Chemical Engineering 24, 2000, 1681 – 1687
- Kumar Vineet and Gupta Raj Kumar Fluid Catalytic Cracking Riser Modeling in Heat Transfer Mode. Chemical Product and Process Modeling Volume 3, Issue 1, 2008, Article 11
- Lee L.-S., Yu S.-W., Cheng C.-T. Fluidized-Bed Catalyst Cracking Regenerator Modelling and Analysis. Chem. Eng. J., 40, 1989, 71.
- McKeen Timothy R., Berry Thomas A., Pugsley Todd S., and Dalai Ajay K., Two-Dimensional Reaction Engineering Model of the Riser Section of a Fluid Catalytic Cracking Unit. Ind. Eng. Chem. Res. 43, 2004, 5571-5581.

- Patience G. S., Chaouki J. F. , Berruti and Wong S. R., Scaling considerations for circulating Fluidized Bed Risers, Powder Technol. 72,1992, 31-39.
- Pinheiro Carla I. C., Fernandes Joana L., Domingues Luís, Alexandre J. S. Chambel, Graca Ines, Oliveira Nuno M. C., Cerqueira Henrique S., Ribeiro Fernando Ramoa, Fluid Catalytic Cracking (FCC) Process Modeling, Simulation,and Control. Ind. Eng. Chem. Res. 2012, 51, 1–29
- Pitualt I., Nevicato, D. Foressier, M. and Bernard, J-R. "Kinetic Model Based on a Molecular Description for Catalytic Cracking of Vacuum Gas Oil". Chem. Eng. Sci, vol. 49, no. 24^{A}, 1994, 4249-4262.
- Pugsley T.S., Bolkan-Kenny Yasemin G. and Berruti Franco, "Computer Simulation of the Performance of Fluid Catalytic Cracking Risers and Downers". Ind. Eng. Chem. Res. 1994, 33, 3043-305
- Theologos, K.N., and Markatos, N.C. Advanced modeling of fluid catalytic cracking riser-type reactors. AIChE J., 39, 1007-1017, 1993
- Weekman V.W., and Nace D.M., Kinetics of Catalytic Cracking Selectivity in Fixed, Moving and Fluid-bed Reactors, AIChE 16, 1970, 397-404