



## MODELLING AND SIMULATION OF COKING IN THE RISER OF AN INDUSTRIAL FLUID CATALYTIC CRACKING (FCC) UNIT

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### ABSTRACT

*Under normal Fluid Catalytic Cracking (FCC) conditions, coke is the most important factor that affects catalyst activity. A pseudo homogeneous two - dimensional (2D) model of an industrial FCC riser is here presented. The FCC riser models of previous researchers were mostly based on the assumption of negligible mass transfer resistance and 1D plug flow. These assumptions undermine the accuracy of the models by over-predicting the riser residence time while under-predicting the reaction time. Mass transfer resistance was incorporated in the reactor model to enhance the accuracy of the results. Finite difference was used to discretise the model equation. This investigation has advanced research into the modeling of FCC riser by predicting catalyst coke content as a function of reaction temperature. The yields of LCO, gasoline, gas and coke that were predicted by the model for industrial risers were 15.54wt%, 49.70wt%, 18.01wt% and 4.90wt% respectively. A feed stock (VGO) conversion of 79.28% was predicted by the model. An optimal operating temperature range of 786K < T < 788K was predicted for the riser.*

*Key words:* FCC, Finite difference, Mass transfer resistance, Catalyst deactivation; Riser models.

### 1. INTRODUCTION

Fluid Catalytic Cracking (FCC) is one of the most profitable processes in oil refineries [2]. It is the major producer of gasoline in refineries and as such it is sometimes referred to as the heart of the refinery. FCC converts vacuum gas oils and heavy feed stocks (molecular weight > 250) from other refinery operations into high octane gasoline, light fuel oils and gases [2].

FCC unit comprises mainly of the riser, the regenerator and the main fractionators. However, the reactor section of FCC units has been the most active area of research in the industry and academia alike as evidenced by the works of previous researchers [1-9]. Most of the reported works were based on assumptions that either over-simplified the obtained models [1-4, 6 and 8-9] or unduly over-complicated it [5 and 7]. For example, Weekman and Nace [1], did not account for coke formation in their model. Hence, the model of the authors could not predict the coking of FCC catalyst. Fernandes *et al.* [2], used a 6-lump, 1D model to simulate the riser of an industrial FCCU. Their model predicted a gasoline yield of 48%. The

temperature, gas and solid phase velocity profiles were also predicted by the authors. However, the assumption of 1D plug flow and negligible mass transfer resistance by the authors over-simplified their models thereby undermining the utility of the predictions.

A 4-lump, 1D model was used by Ahari *et al.* [3]. The authors were able to capture the temperature drop along the riser and they predicted a gasoline yield of 45%. The major setback of their model was also the assumption of negligible gas phase dispersion. A 5-lump reaction scheme was used by Alsabei [4]. The author also based his investigation on negligible dispersion which contradicts the basic principles of heterogeneous catalysis, especially, for porous catalysts such as the molecular sieve type commonly used as FCC catalysts. You and Zhu [6], used a 1D, 4-lump model to predict the products yield of FCCU riser, the average gas density, solid and gas phase velocity profiles all as a function of the riser height. The authors predicted a gasoline yield of 45%. However, non-incorporation of mass transfer resistance in their model over-simplified the model. A

4-lump, 1D scheme was also used by Heydari *et al.* [8], to model an industrial riser. The yield of gasoline along the riser under varying conditions of temperature and catalyst-to-oil ratio was predicted by the authors.

In reality the riser is a 3D reactor. Simplifying the geometry to 1D is tantamount to predicting products yield just along the axis of the reactor. However turbulent the flow in the riser may be, a 1D model cannot adequately represent the entire geometry of the reactor because it does not account for wall effects. Even though some of these cited works were able to predict quite reasonable gasoline yield value (45 – 48%), the models are very limited in utility due to their wholesome empirical nature. They offer little understanding of the process, hence restricting scale up to within the process parameters range investigated.

Models of higher dimensionality have also been used by other authors [5, 7 and 9]. Souza *et al.* [9], used a 2D hydrodynamic, 6-lump model to simulate an industrial riser. The model predicted a gasoline yield of 48%. The authors also neglected mass transfer resistance in their model. Gupta [5], and Lopes *et al.* [7], used 3D models in their investigations. Gupta [5], used a mechanistic approach involving 50 lumps (pseudo species) to model an industrial FCCU. Lopes *et al.* [7], on the other hand, used a 4-lump reaction scheme to investigate the effects of various exit configurations of the riser on the hydrodynamics of the reactor as well as the yield of gasoline. They found that the T-shape exit configuration enhanced the yield of gasoline owing to enhanced solid (catalyst) reflux. However, 3D models are very complex and unwieldy. They also have high costs of computation. In all the models aforementioned, the authors did not simulate the catalyst coke content thereby leaving room for more work to be done in that respect.

This work sought to improve on the existing reports by incorporating a semi-empirical approach via integration of mass transfer resistance scheme, thereby accounting for the lower experimental gasoline yield and longer reaction time when compared to the over-simplified models. This work also explains in real engineering terms the empirical results earlier models reported as functions of vessel geometry. This is to involve a model more amenable to extrapolation. A two-dimensional (2D) quasi-steady state model of an industrial riser is here presented. In this work, a five-lump reaction scheme was used to model the FCC reactions. This investigation has also

advanced the works of previous researchers in this field by simulating the catalyst coke content with a view to predicting the operating conditions that will minimize the coking of FCC catalyst thereby reducing the cost of regeneration of the coked catalyst.

## 2. MATERIALS AND METHODS

The FCCU reactor was modeled in this work using MATLAB (R2009a) on a Compaq HP CQ61 laptop.

The following assumptions were made in the development of the model:

1. Pseudo homogenous two-dimensional transport with axial and radial gradients.
2. The catalyst and gas are at thermal equilibrium
3. Hydrocarbon feed comes into contact with the hot catalyst coming from the regenerator and instantly vaporizes [5].
4. There is no heat loss from the riser, and the temperature of the reaction mixture (hydrocarbon vapors and catalyst) falls only because of the endothermicity of the cracking reactions [5].
5. The riser dynamic is fast enough to justify a quasi-steady state model.

Figure 1 depicts the five-lump reaction scheme that was used in this investigation.

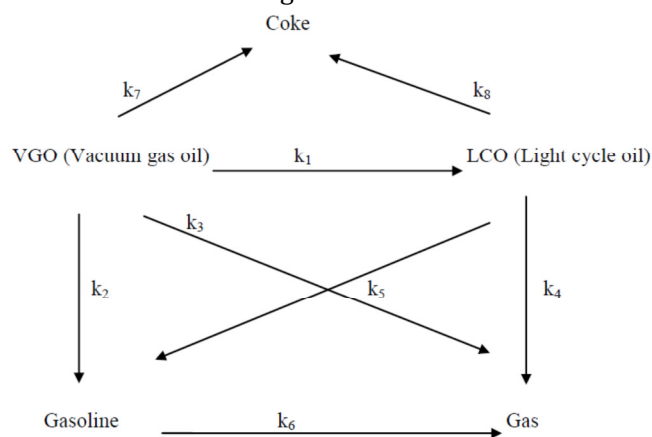


Figure 1: Five-lump model [10]

In Figure 1,  $k_j$  is the rate constant of reaction  $j$  in  $s^{-1}$  where  $j=1, 2, \dots, 8$ .

### 2.1 Model rate equation

In the five-lump model given in Figure 1, the eight reactions of the model are taken to follow first order kinetics as follows:

$$r_j = \frac{ac_i}{\left(\frac{1}{k_g} + \left(\frac{1}{\eta k_j}\right)\right)} \quad j = 1, \dots, 8 \quad (1)$$

In Equation (1),  $r_j$  is the rate of reaction of the  $j$ th reaction in  $kg$  species  $(kg \text{ catalyst})^{-1}s^{-1}$ ,  $a$  represents

catalyst exponential activity decay given by Equation (2),  $c_i$  = species concentration (weight fraction),  $k_g$  = mass transfer coefficient of reactant in m/s,  $\eta$  = particle effectiveness factor which is given by Equation (3),  $k_j$  = reaction rate constant in  $s^{-1}$ .

$$a = \exp(-k_d c_{coke}) \quad (2)$$

Where  $k_d$  is the catalyst decay constant and has a value of 8.2 [10].

$$\eta = \frac{3}{\varphi} \left( \frac{1}{\tanh \varphi} - \frac{1}{\varphi} \right) \quad (3)$$

$\varphi$  is the Thiele modulus.

$$\varphi = R \left( \frac{k_j}{D_e} \right)^{\frac{1}{2}} \quad (4)$$

$D_e$  = effective diffusivity in  $m^2/s$ . Equation (1) is the model rate equation which incorporates mass transfer resistance terms,  $k_g$  and  $\eta$ . Equation (1) reverts to the classical first order rate equation when  $\frac{1}{k_g} = 0, \eta = 1$ .

(Previous researchers [10] used the classical first order rate equation:  $r_j = k_j c_i$ ). The particle effectiveness factor,  $\eta$  expressed by Equation (3) is the ratio of the reaction rate when there is diffusion resistance to the rate when there is no diffusion resistance. It is a direct measure of the extent to which diffusion resistance reduces the rate of chemical reactions in solid catalysis and it is a function of Thiele modulus. Equation (4) holds for spherical particles. Thiele modulus,  $\varphi$  is the ratio of intrinsic reaction rate to diffusion rate and as such Equation (4) provides a yardstick for determining the rate-determining step in heterogeneous reactions involving solid catalysts.

The basic parameters to be determined in Equations (1)-(4) are  $D_e$  and  $k_g$ .  $D_e$  was estimated from empirical correlations in literature [11], while  $k_g$  was estimated from Sherwood number for gases [12].

**2.2 Riser reactor model equations**

Figure 2 depicts the 2D riser reactor, while the control volume used in deriving the model equations from conservation laws is shown in Figure 3 [11].

**2.2.1 Continuity equation**

The component continuity equation for the model is as given in Equation (5):

$$D_{zi} \frac{\partial^2 c_i}{\partial z^2} + D_{ri} \left( \frac{\partial^2 c_i}{\partial r^2} + \frac{1}{r} \frac{\partial c_i}{\partial r} \right) - \frac{\partial (u c_i)}{\partial z} - \rho_B (-r_i) = 0 \quad (5)$$

Where the superficial velocity,  $u$  is given by Equation (6).

$$u = \frac{q}{A_c}, m^3(fluid)s^{-1}m^{-2}(vessel) \quad (6)$$

$q$  is the volumetric flow rate of the gas through interparticle bed voidage,  $m^3(fluid)s^{-1}$ ,  $D_z$  and  $D_r$  are effective diffusivities in  $m^2(fluid)m^{-1}(vessel)s^{-1}$ ,  $(-r_i)$  is in  $kg\ species\ kg^{-1}(catalyst)\ s^{-1}$ ,  $A_c$  is the cross-sectional area of the riser in  $m^2$ ,  $c_i$  is concentration of species,  $i$  in weight fraction,  $\rho_B$  is the density of the bed in  $kg/m^3$ ,  $r$  and  $z$  are the radial and axial distances respectively in  $m$ .

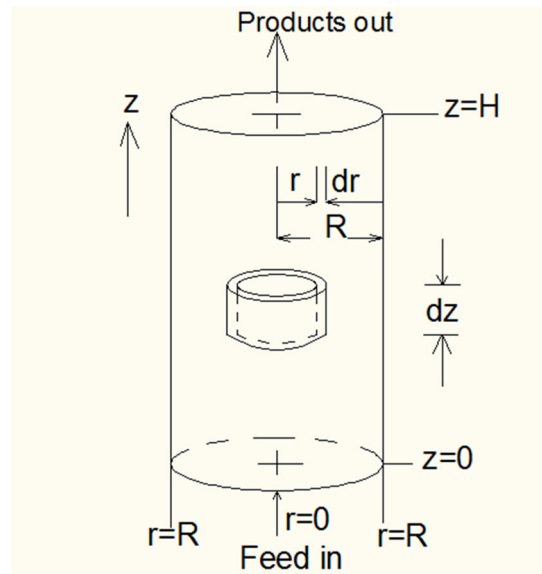


Figure 2: Control volume in riser reactor

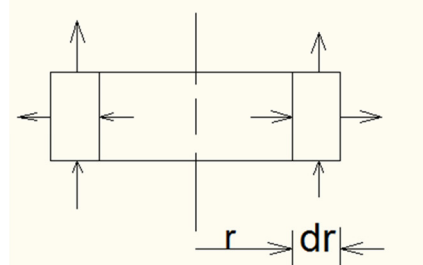


Figure 3: Control volume

**2.2.2 Riser hydrodynamic model**

The numerical value of the catalyst slip factor (the ratio of the gas interstitial velocity to the average particle velocity) can be predicted from Equation (7) [3]:

$$\psi = \frac{u_0}{\epsilon v_p} = 1 + \frac{5.6}{Fr} + 0.47 Fr_t^{0.47} \quad (7)$$

In (7),  $Fr$  is the Froude number and  $Fr_t$  is the Froude number at terminal velocity,  $\psi$  is the catalyst slip factor,  $\epsilon$  is the average voidage of the reactor and  $v_p$  is the average particle velocity in the riser in  $m/s$ .

$$Fr = \frac{u_0}{(gD)^{0.5}} \quad (8)$$

Here,  $g$  is the acceleration due to gravity ( $m^2/s$ ),  $D$  is the diameter of the riser in m. The average particle velocity in the riser,  $v_p$  is given by Equation (9).

$$v_p = \frac{G_s}{\rho_s(1 - \varepsilon)} \quad (9)$$

$G_s$  is the catalyst mass flux in  $kg/m^2.s$ ,  $\rho_s$  is density of solid in  $kg/m^3$ . The expression for the average voidage in terms of the solid mass flux, superficial gas velocity, riser diameter and catalyst physical properties was derived from Equations (7) and (9). Equation (10) gives the average voidage of the reactor.

$$\varepsilon = 1 - \frac{G_s \psi}{u_0 \rho_s + G_s \psi} \quad (10)$$

### 2.2.3 Energy balance

Equation (11) is the model energy balance.

$$k_z \frac{\partial^2 T}{\partial z^2} + k_r \left( \frac{\partial^2 T}{\partial r^2} + \frac{1}{r} \frac{\partial T}{\partial r} \right) - G_s c_p \frac{\partial T}{\partial z} + \rho_B \sum_{i=1}^8 (-r_i)(-\Delta H_{Ri}) = 0 \quad (11)$$

In (11),  $k_z$  and  $k_r$  are the effective thermal conductivities in  $W/m.K$ ,  $c_p$  is specific heat capacity in  $J/kg.K$ ,  $\Delta H_{Ri}$  is enthalpy of cracking of species  $i$  in  $J/kg$ . The coupling between the riser and the regenerator is expressed in the model by Equation (12).

$$F_{cat} c_{pcat}(T_0 - T_{cat}) + F_f c_{pfl}(T_{vap} - T_f) + F_f c_{pfv}(T_0 - T_{vap}) + F_f \Delta H_{vap} = 0 \quad (12)$$

$F_{cat}$  is catalyst flow rate in  $kg/s$ ,  $F_f$  is feed flow rate in  $kg/s$ ,  $c_{pcat}$  is the specific heat capacity of the catalyst in  $kJ/kg.K$ ,  $c_{pfl}$  is the specific heat capacity of liquid feed in  $J/kg.K$ ,  $c_{pfv}$  is the specific heat capacity of vapor feed in  $J/kg.K$ ,  $T_0$  is inlet temperature of riser in  $K$ ,  $T_{cat}$  is inlet temperature of catalyst in  $K$ ,  $T_f$  is feed inlet temperature in  $K$ ,  $T_{vap}$  is feed vaporization temperature in  $K$  and  $\Delta H_{vap}$  is enthalpy of vaporization of the feed in  $J/kg$ . The governing equations, Equations (5) and (11) were expressed in a general, normalized form as follows:

$$\alpha \left( \frac{\partial^2 \sigma}{\partial r^{*2}} + \frac{1}{r^*} \frac{\partial \sigma}{\partial r^*} \right) + \beta \frac{\partial^2 \sigma}{\partial z^{*2}} + \gamma \frac{\partial \sigma}{\partial z^*} + \lambda(-r_i) = 0 \quad (13)$$

Here,  $\sigma = c_i/c_0$  or  $T/T_0$ ,  $r^* = r/R$ ,  $z^* = z/H$ . The coefficients in Equation (13) are given by the following expressions:

$$\alpha_1 = \alpha_2 = 1, \beta_1 = \frac{R^2 D_z}{H^2 D_r}, \beta_2 = \frac{R^2 k_z}{H^2 k_r}, \gamma_1 = \frac{-UR^2}{HD_r}, \gamma_2 = \frac{-Gc_p R^2}{Hk_r}, \lambda_1 = \frac{R^2 \rho_B}{D_r c_0}, \lambda_2 = \frac{R^2 \rho_B}{k_r T_0} \quad (14)$$

Subscripts 1 and 2 in the coefficients in Equation (14) correspond to the continuity equation and energy balance respectively.

Boundary conditions:

$$@ z^* = 0, 0 < r^* < 1 \text{ (inlet): } \sigma_{vgo} = \sigma_T = 1, \sigma_{lco} = \sigma_{gasoline} = \sigma_{gas} = \sigma_{coke} = 0$$

$$@ z^* = 1, 0 < r^* < 1 \text{ (outlet): } \frac{\partial \sigma}{\partial z^*} = 0$$

$$@ r^* = 0, 0 < z^* < 1: \frac{\partial \sigma}{\partial r^*} = 0$$

$$@ r^* = 1, 0 < z^* < 1: \frac{\partial \sigma}{\partial r^*} = 0 \quad (15)$$

Finite difference numerical scheme was used to discretize the governing equation; Equation (13) that was developed for the model. A code was written in MATLAB to solve the model equations. Data obtained from open literature [2] and from Kaduna Refinery and Petrochemicals Company (KRPC) were used to validate the model results. Thereafter, catalyst coke content was simulated to predict the coke content of the FCC catalyst for a selected reactor temperature range.

### 2.3 Model data

The data used for the simulation are as given in Tables 1-5

Table 1: Kinetic constants for five-lump model

Reaction number	Kinetic constant, $s^{-1}$
1	1.90
2	7.50
3	1.50
4	0.00
5	1.00
6	0.30
7	0.21
8	0.50

Source [10]

Table 2: Enthalpies of cracking

S/N	Cracking reaction	Enthalpy, $kJ/kg$
1	VGO to LCO	80
2	VGO to gasoline	195
3	VGO to gas	670
4	LCO to gas	-
5	LCO to gasoline	180
6	Gasoline to gas	530
7	VGO to coke	745
8	LCO to coke	600

Source [3]

Table 3: Molecular weights and heat capacities

S/N	Species	Molecular weight, kg/kmol	Specific heat, kJ/kg.K
1	VGO	333.0	2.67 (liquid), 3.30 (gas)
2	LCO	300.0	3.30
3	Gasoline	106.7	3.30
4	Gas	40.0	3.30
5	Coke	14.4	1.087

Source [3]

Table 4: Gas oil properties

Property	Value	Source
Specific gravity	0.89-0.93	[5]
Viscosity	$1.4 \times 10^{-5} \text{N.s/m}^2$	[3]
Vaporization temperature	698K	[3]
Enthalpy of vaporization	190kJ/kg	[3]

Table 5: Model parameters (Source: KRPC Plant data)

S/N	Parameter	Value	Source
1	Reactor inlet temperature, $T_0$ (K)	791	KRPC Plant data
2	Feed inlet temperature, $T_f$ (K)	613	KRPC Plant data
3	Catalyst inlet temperature, $T_{cat}$ (K)	927	KRPC Plant data
4	Specific heat capacity (liquid feed), $cp_{fl}$ (J/kg-K)	2.67e3	[3]
5	Specific heat capacity (vapor feed), $cp_{vf}$ (J/kg-K)	3.30e3	[3]
6	Specific heat capacity (catalyst), $cp_{cat}$ (J/kg-K)	1.09e3	[3]
7	Feed vaporization temperature, $T_{vap}$ (K)	698	KRPC Plant data
8	Enthalpy of vaporization, $\Delta H_{vap}$ (J/kg)	190e3	[3]
9	Density (solid catalyst), $\rho_s$ (kg/m <sup>3</sup> )	1250	KRPC Plant data
10	Catalyst velocity, $U_c$ (m/s)	5	[5]
11	Gas superficial velocity, $U$ (m/s)	18	KRPC Plant data
12	Slip factor, $\psi$	2	KRPC Plant data
13	Feed flow rate, $F_f$ (kg/s)	35.5	KRPC Plant data
14	Riser diameter, $D_R$ (m)	1.146	KRPC Plant data
15	Riser height, $H$ (m)	25	KRPC Plant data

S/N	Parameter	Value	Source
16	Pore diameter, $P_d$ (m)	2.00e-9	KRPC Plant data
17	Particle diameter, $D_p$ (m)	60e-6	KRPC Plant data
18	Gas average density $\rho_g$ (kg/m <sup>3</sup> )	0.92	KRPC Plant data
19	Gas average viscosity $\mu_g$ (Pa.s <sup>-1</sup> )	1.40e-5	[3]
20	Riser pressure, $P$ (atm)	2.94	KRPC Plant data
21	Particle tortuosity, $\tau_p$	7	[11]

### 3. RESULTS AND DISCUSSION

The results obtained at the end of the investigation were presented as shown in Figures 4, 5 and 6. The predicted yields of LCO, gasoline, gas and coke as depicted in Figure 4 are 15.54wt%, 49.70wt%, 18.01wt% and 4.90wt% respectively. These values compare favorably well with literature and plant (KRPC) data (Table 5). Figure 4 shows that the products of FCC reactions are formed within the first 2m of the reactor (the reaction zone). This is because feed vaporization occurs in the reaction zone followed by cracking which occurs instantaneously as the vapor feed contacts the hot catalyst entering the riser from the regenerator. Also, in Figure 4, it can be seen that after reaching the peak value, the yield of each of the products remains constant throughout the remaining length of the riser. This is because the residence time for industrial risers is set at 2s so that the gaseous products are separated from the catalyst soon enough to avoid over-cracking of gasoline [5].

Figure 5 presents the predicted conversion of VGO as a function of reactor height. A conversion of 79.28% was predicted by the model. In Figure 5, VGO conversion is observed to occur within the first 2m of the reactor which corresponds to the reaction zone of the reactor. The reaction zone is also the zone within which the products are formed as depicted in Figure 4. The other zones of the reactor are the middle and upper zones. The middle zone is the region of gasoline over-cracking for maximum gas production when gas becomes the key product.

The validation of model results with KRPC and literature data is as shown in Table 5. The deviation of the model predictions from KRPC plant data were computed and presented in the last column of Table 5. The industrial riser model here presented compare favorably well with KRPC plant data and literature results because the maximum deviation of the predictions is 3.54%. This value is less than the error limit of 5%.

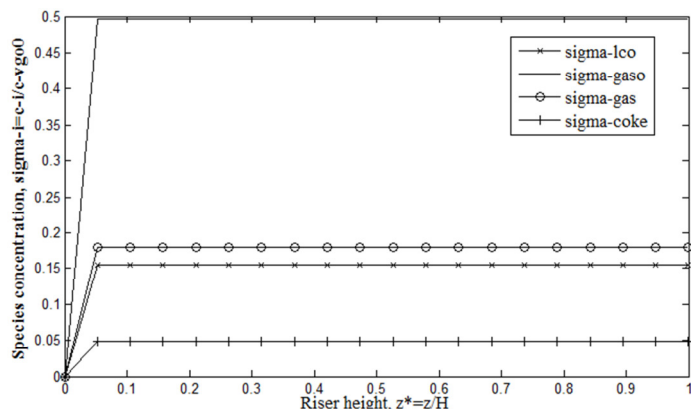


Figure 4: FCC products concentration (wt %) along riser height

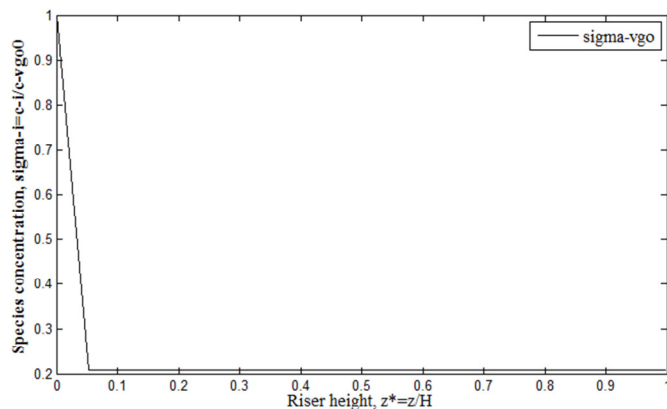


Figure 5: Feedstock (VGO) conversion along FCC riser height

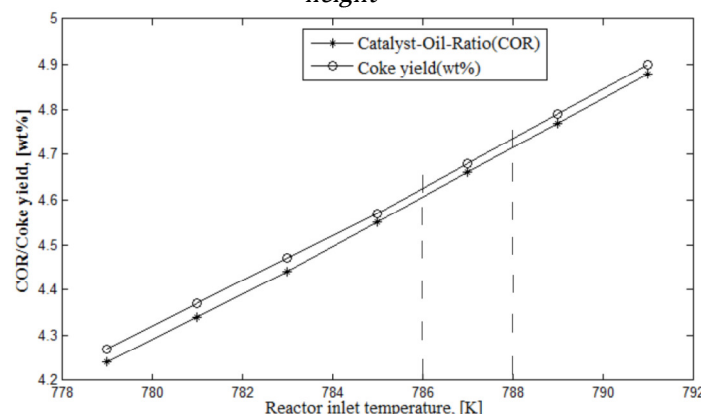


Figure 6: Model result for simulation of FCC catalyst coking

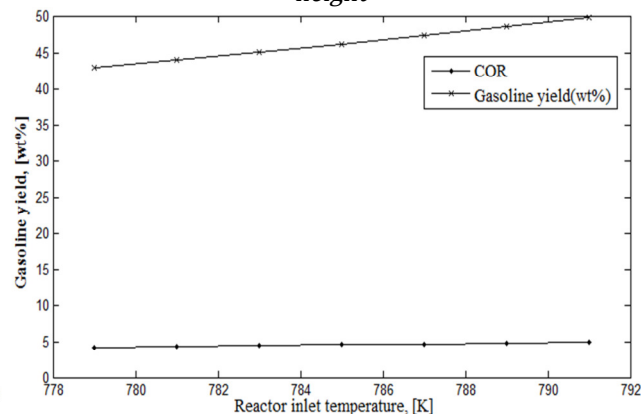


Figure 7: Variation of COR and gasoline yield with reaction temperature

Table 5: Validation of model results with plant data

Species	Conv./Yield, wt% (KRPC plant data)	Conv./Yield, wt% [2]	Conv./Yield, wt% (Model)	% Dev. from Plant data
VGO	80.00	78.00	79.28	0.90
LCO	15.15	10.00	15.54	2.57
Gasoline	50.00	48.00	49.70	0.60
Gas	17.88	18.00	18.01	0.73
Coke	5.08	5.00	4.90	3.54

Coke on catalyst was simulated using the validated model. The result was presented as a plot of catalyst coke content as a function of reactor temperature as shown in Figure 6. In Figure 6, it can be seen that the reactor temperature increases monotonously as the catalyst-to-oil ratio (COR) increases. This is because the enthalpy of the incoming catalyst from the regenerator determines the temperature at the inlet and the outlet of the riser.

Three critical temperature regimes were identified from the plot in Figure 6. These are:

i. Low operating temperature regime ( $T < 786K$ ): If the riser is operated in this regime (lower region of

the graph), the reactions will quench. Hence, operation in this regime is not advisable.

ii. Optimal operating temperature regime ( $786K < T < 788K$ ): In this temperature range, COR and catalyst coke content profiles taper towards each other as shown in Figure 6. This is the regime of optimal riser operation (without excessive coking). Plant operation at reduced coking rate will reduce the cost of catalyst regeneration which in turn increases plant profitability.

iii. High operating temperature regime ( $T > 788K$ ): In this temperature zone, the two curves tend to diverge from each other again symbolizing excessive coking of the catalyst. Unit operation in this temperature range is also not advisable because it leads to excessive coking and gas production at the expense of the most economical product (gasoline).

Figure 7 depicts the variation of COR and the yield of gasoline with reactor temperature. The average yield of gasoline within the proposed optimum temperature range ( $786K < T < 788K$ ) is 47.33%; this value falls within the range of gasoline yield predicted in



literature [2 and 10] and obtainable from KRPC (45%-50%).

**4. CONCLUSIONS**

At the end of this investigation, the yields of LCO, gasoline, gas and coke predicted by the model for industrial risers are 15.54wt%, 49.70wt%, 18.01wt% and 4.90wt% respectively. A VGO conversion of 79.28% was predicted by the model. It can be inferred from the results of this investigation that an operating temperature range of 786K<T<788K and a catalyst to oil ratio (COR) range of 4.60-4.71 are optimal for FCC. Amongst the operational objectives of FCC unit are plant operation at reduced coking rate (coke is expensive to burn off the catalyst) and that gas production should not be in excess for gas is less valuable and yet expensive to compress. FCC riser operation within the optimal temperature range ensures that these economic objectives are achieved by minimizing the coking of the catalyst while ensuring that the desired product (gasoline) is not over-cracked to gas. The results of this investigation also showed that feed stock conversion and product formation occur within the first 2m of the reactor.

**NOMENCLATURE**

a	Catalyst activity
$c_i$	Species concentration (weight fraction)
$c_p$	Specific heat capacity (J/kg-K)
$d_{AB}$	Collision diameter (m)
$D_{AB}$	Molecular diffusivity (m/s <sup>2</sup> )
$D_e$	Effective diffusivity (m/s <sup>2</sup> )
$D_k$	Knudsen diffusivity (m/s <sup>2</sup> )
$D_p$	Particle diameter (m)
$D^*$	Overall diffusivity (m/s <sup>2</sup> )
$F_i$	Flow rate of species $i$ (kg/s)
$G_s$	Catalyst mass flux (kg/m <sup>2</sup> .s)
$\Delta H_{Ri}$	Enthalpy of cracking of species $i$ (kJ/kg)
$\Delta H_{vap}$	Enthalpy of vaporization (kJ/kg)
P	Pressure (atm)
$r_e$	Average pore radius (m)
$r_i$	Species reaction rate (kg species (kg catalyst) <sup>-1</sup> .s <sup>-1</sup> )
k	Reaction rate constant (s <sup>-1</sup> )
$k_d$	Catalyst decay constant
$k_r, k_z$	Effective thermal conductivity (W/m.K)
$k_g$	Mass transfer coefficient (m/s)
t	Time (s)
X	Conversion
$M_i$	Molecular weight species $i$ (kg/kmol)
m	Node number in the horizontal direction
T	Temperature (K)
R	Radius (m)
$\Delta r$	Radial spatial interval (m)
n	Node number in the vertical direction
$N_A$	Molar flux (kmol/m <sup>2</sup> .s)
$N_{Re}$	Particle Reynolds number
$N_{sc}$	Schmidt number
$N_{sh}$	Sherwood number

$N_r$	Number of divisions in radial direction
$N_z$	Number of divisions in axial direction
V	Reactor volume (m <sup>3</sup> )
$v_{ij}$	Stoichiometric coefficient
$v_p$	Average particle velocity (m/s)
H	Reactor height (m)
u	Superficial velocity (m/s)
q	Volumetric flow rate (m <sup>3</sup> /s)
$A_c$	Cross-sectional area (m <sup>2</sup> )
$F_r$	Froude number
$\Delta z$	Axial spatial interval (m)

*Greek letters:*

$\alpha'$	Decay function rate constant
$\alpha$	Normalized parameter
$\beta$	Normalized parameter
$\gamma$	Normalized parameter
$\delta$	Decay function constant
$\epsilon$	Porosity
$\eta$	Particle effectiveness factor
$\eta_0$	Particle overall effectiveness factor
$\lambda$	Normalized parameter
$\mu$	Viscosity (Pa.s <sup>-1</sup> )
$\pi$	Pi
$\phi$	Thiele modulus
$\psi$	Slip factor
$\rho$	Density (kg/m <sup>3</sup> )
$\sigma$	Normalized variable
$\tau$	Tortuosity
$\Omega_D$	Collision integral

*Subscripts:*

$i$	Species number
$j$	Reaction number

*Abbreviations:*

KRPC	Kaduna Refinery & Petrochemicals Company Ltd
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