Control-Relevant Dynamic Model Development and Validation for an Industrial Fluid Catalytic Cracking Unit

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Abstract- **In this paper an industrial Fluid catalytic Cracking unit was discretized into four sub-systems (vapourizer, riser, separator catalyst regenerator) to adequately capture its control-relevant features with model representations. While the vapourizer was modelled as a steady state heat exchanger, the regenerator as a continuously stirred tank reactor, the riser was modelled as a tubular reactor with plug flow in space and time, yielding a set of partial differentials, differential-algebraic equations. The riser partial differential equations were then reduced by order one, using the method of lines with eighteen internal nodes and two boundary nodes. The equations, along with constitutive relations that were developed based on hydrodynamics and catalyst activity decline considerations were solved following a sequential-modular approach with codes that were developed in this study and implemented in MatLab. Model predictions in comparison with plant data: gasoline yield (47.74%/45.90%): Light gases yield (25.65%/26.60%): Coke yield (5.38%,5.1%): Unconverted gasoil (21.56%/22.40%); riser exit temperature(527ºC/524ºC); regenerator temperature (774ºC/743ºC);coke on regenerated catalyst ,mole(0.046%/0.05%),oxygen in regenerator flue gas ,mole(2.84%,3.0%) show reasonable agreement and suggest the suitability of the model for simulation and control evaluation of the FCC unit. The novel nodal discretization of the riser partial equations and at each node allows for the calculation of catalyst activity decline as a function of space, time and temperature. compensates for the over-simplifications in the lumped parameter kinetics, thereby approximating the single event kinetics. More so, the sequential-modular solver that was developed for the solution of the large set of differential algebraic equations is in-house, home-grown and novel.**

Keywords: **Process Model, Lumping, Voidage, Hydrodynamics, Catalytic Cracking**

1: INTRODUCTION

Fluid catalytic cracking (FCC)is the conversion of high-boiling, low- value feeds such as vacuum gasoil to low-boiling, high -value products such as liquified petroleum gas and gasoline in a fluidized catalyst bed reactor. It is commonly referred to as the heart of a refinery, in the light of its role as the highest contributor to the gasoline pool. However, owing to its complexity that is occasioned by several themes and phenomena such as feed vapourization and atomisation, simultaneous heat as mass transfer, hydrodynamics, reaction kinetics and catalyst deactivation, holistic understanding of all aspects of FCC is lacking. A direct approach to improve the understanding of FCC behaviour is to study various operational scenarios, using an industrial FCC unit or a pilot plant. Such option is often not available due to stringent policies, cost and safety considerations among others, prompting, modelling of the FCC to be attractive. In this regard,a one-dimensional ,steady state, plug flow model of a fluid catalytic cracking riser was reported in Babatope et al. (2013).In Josiah et al (2014), however, the plug flow scheme was restricted to the riser that was modelled at steady state while the regenerator was modelled as a continuously stirred tank reactor(CSTR).An Eulerian multi-phase approach to the modelling of the fluid catalytic cracking unit was presented in Ahmed and Atega (2016) while in John et al.(2017) the same approach as in Babatope et al.(2013) was followed. Elsewhere in the literature Dagde and Puyate (2012), Dagde and Akpa (2014), Du et al (2014), Olafadehan et al. (2018, 2019) presented process models of the fluid catalytic cracker riser and regenerator that relied heavily on empirical correlations. Concerning kinetic model that describe the chemistry of the FCC reactions, there are primarily two lines of argument, one in favour of single event models and the other in favour of lumped parameter models. Due to the overwhelming number of molecular species that are present in vacuum gas oil, the feed to the FCC unit, researchers appear to have largely embraced the later due to its simplicity in application. In Baudrez et al (2010) the 4-lump kinetic scheme was adopted as a kinetic descriptor. Several other authors, including Josiah et. al (2014), Dagde and Akpa (2014), Ahsan (2015), Ahmed and Atega (2016) Polgar and Somayeh (2018) and Guan et al (2019) adopted the 4-lump model to FCC modelling stuidies. However, the adoption of the 5-lump kinetic model was reported in Dagde and Puyate (2012) while Du et al (2014) and Olafadehan et al (2019) adopted 6-lump kinetic model. Other kinetic lump models such as the 11-lump and the 17-lump models were adopted in. Yang et al. (2016) and Sigh et al (2017) respectively. A comparative study of 4-lump and 3-lump superiority in which the 4-lump was favoured was reported in Cristi (2015). Cristi (2015) argues that from the standpoint of higher gasoline yield in simulation studies, the 4-lump model is a better option than the 3-lump. Another important theme that is associated with FCC studies is deactivation and activity decline. Catalyst deactivation following coke deposition plays a major role in FCC operations and it is accounted for through the deployment of activity decline models in the modelling framework . Coke-on –catalyst and the catalyst time on stream are the two types of such models that have been reported in the literature. In Babatope et al. (2013), John et al (2017) and Guan et al. (2019) catalyst deactivation was modelled by the coke-on-catalyst representation. While there is no strong reason for this choice apart from convenience my analysis is that for modelling efforts that focus only on the riser, the coke on catalyst

model is a flawed on, from the point of absence of a regenerator model to account for coke-burn. The time on stream model accounts for the residence time of the catalyst in the regenerator where the coke-burn reactions take place. Perhaps it is on this strength that it has been overwhelmingly adopted in recent FCC simulation such as the work of Ahsan (2015), Ahmed and Atega (2016) ,Yang et al.(2016) , Sigh et al (2017) , Porgar and Somayeh(2018), Olafadehan et al.(2019), and Guan et al (2019)

Concerning feed vaporization during FCC operations, there seems to be agreement across board on the validity of instantaneous vaporization of gas oil feed at the riser entrance. Dagde and Puyate (2012), Babatope et al. (2013), Du et al (2014), Josiah et al (2014), Ahsan(2015), Ahmed and Atega(2016) ,Yang et al.(2016) , Sigh et al (2017) , John et al (2017) ,Porgar and Somayeh(2018), Olafadehan et al.(2019), Guan et al (2019) , and Guan et al.(2019), are all in tandem with the instantaneous vaporization proposition. Moreover, Olafadehan et al (2019) has shown that there is no significant difference in simulated results between one dimensional transient model and instantaneous vaporization as descriptors of FCC feed vaporization at the riser entrance. From the point of feed vaporization up to the riser termination zone, gas phase transport, solid phase transport and the interaction between the gas and solid phase, commonly referred to as hydrodynamics, is an important factor in FCC studies and operations. Multi-phase particle-in cell (MP-PIC) models that are based on computational fluid dynamics (CFD) and its variants are basically employed to describe riser hydrodynamics. Such models are either of the Eulerian-Eulerian formulation or the Eulerian-Lagrangian formulation in which the fluid phase is given an Eulerian representation while the particle (solid phase) is given a Lagrangian representation in the momentum balance. The work of Alveris-Castro (2015), Xu et al (2018), Dutta et al (2018) and Yang and Wang (2020) are in this category that are said to be highly efficient in mimicking gas-particle motion in the FCC riser or downer. However, they are three or two-dimensional two-phase models whose solution is not tractable and of heavy computational load. While the models are suitable for standalone hydrodynamic studies, they do not support control and optimization studies. this perhaps, is the reason that births the plethora of research articles in which voidage (the mass fraction of particle that is present in a particle-in-cell sample) is considered as an adequate representation of gas-particle interaction in FCC modelling. In this regard, Josiah et al (2019) gave an algebraic expression for the computation of voidage along the riser height, based on primary volume fraction definition. However, in Olafadehan et al (2019), the algebraic expression given for voidage relies heavily on empirical correlations and dimensionless quantities such as Froude's number and Reynolds number which in turn are functions of several other variables. The purpose of this paper is to present and validate a dynamic model of the fluid catalytic cracking process that is suitable for evaluating control structures, controller design as well as controller performance testing and evaluation

2: METHOD

A two -step procedure consisting of model development, model solution and model validation was followed in the course of this study. The details are as presented in section 2.1 and section 2.2 respectively

2.1: Model Development

2.1.1: Kinetic Model

The endothermic reactions that take place in the riser are very significant in the modelling of the FCC process. In this paper, the four-lump model of Lee et al. (1989), shown in figure 1, was adopted as a descriptor of FCCU riser reactions. As inferred from the four-lump scheme, gasoil when catalytically cracked gives rise to gasoline light gases and coke which is deposited on the catalyst. The temperature in the riser also favours the cracking of gasoline to form light gases and coke.

While the cracking reactions follow second kinetics with respect to gasoil, it is first order with respect to gasoline. Representing the mass fraction of gasoil by y_1, those of gasoline by y_2, light gases by y_3 and coke mass fraction by y_4, the following rate expressions subsist:

$$(-r_{gasoil}) = (k_{12} + k_{13} + k_{14}) y_1^{\,2} \varphi \quad (1)$$

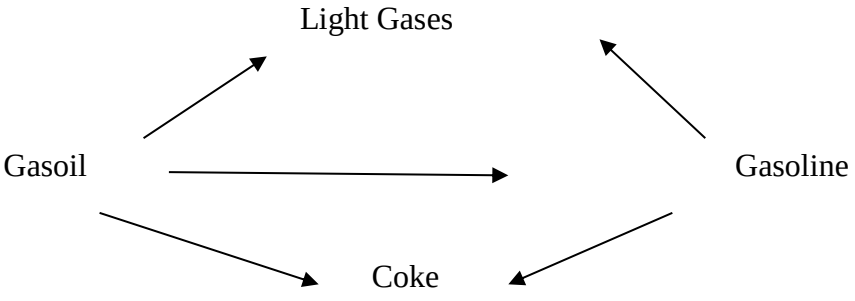
$$(-r_{gasolinel}) = -(k_{12} y_1^{\,2} - (k_{23} + k_{24}) y_2) \varphi \quad (2)$$

$$(-r_{lightgases}) = -(k_{13} y_1^{\,2} + k_{23} y_2) \varphi \quad (3)$$

$$(-r_{coke}) = -(k_{14} y_1^{\,2} + k_{24} y_2) \varphi \quad (4)$$

The catalyst activity decline over time, φ, is related to the flow space time, t_v and the riser temperature, T_{rx} as given in Josiah et al (2019) and used here in that form as follows.

$$\varphi = \exp(-a t_v \exp(-\frac{E}{R T_{rx}})) \quad (5)$$



2.1 .2: Process Models

A modular approach in which the FCC unit is discretized into four sub-systems (see Figure 2) namely: vapourizer, riser, separator and regenerator was employed in this paper.

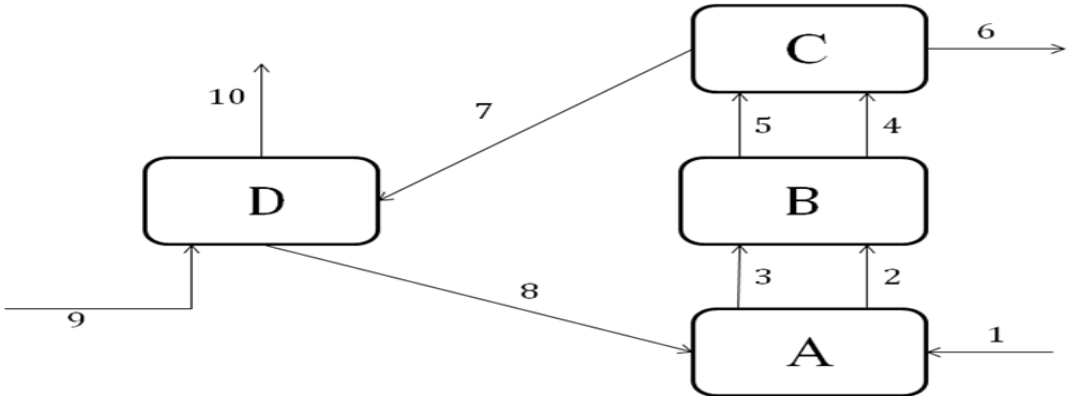


Figure 2. Sub-systems conceptualization of Fluid catalytic cracking unit (FCCU)
A: Vaporization zone, B: Riser, C: Separator, D: Regenerator
1. Feed Gas Oil Stream, 2. Evaporated Gas Oil Leaving the Vaporizer, 3. Catalyst Stream Leaving the Vaporizer
4. Gas Oil Vapour Leaving the Riser, 5. Spent Catalyst Stream Leaving the Riser,
6. Cracked Gas Oil Product 7. Spent Catalyst Stream Leaving the Separator,
8. Regenerated Catalyst Stream, 9. Air Stream entering the Regenerator, 10. Spent Air Leaving the Regenerator

2.1.2.1: Vaporizer Sub-System Model
The region at which the hot regenerated catalyst meets with the feed gas oil is conceived of as a vapourizer. Here, the catalyst raises the feed temperature to its boiling point and provides the heat of reaction that is required to sustain the endothermic riser reactions. At this point the feed stream and the catalyst eventually attain an equilibrium temperature, T_A. A steady-state energy balance around this sub-system serves to establish this gas-particle temperature.
Assuming that the hold-up in the sub-system is sufficient to ensure thermal equilibration between the streams, then the temperatures of these streams under no-slip conditions will be the same (T_A) , based on steady state energy balance given in equation 6

$$F_{gR}\left(C_{pl}(T_g^{/}-T_1)+\Delta\underline{H}^{vap}+C_{pg}(T_A-T_g^{/})\right)=F_cC_{pc}(T_D-T_A) \quad (6)$$

$T_{A,}$ may, therefore, be determined from equation (6) as,

$$T_A=\frac{F_cC_{Pc}T_D+F_{gR}(C_{Pl}T_1-\Delta C_P^{Av}T_g^{/}-\Delta\underline{H}^{vap})}{F_cC_{Pc}+F_{gR}C_{Pg}} \quad (7)$$

with $\Delta C_P^{Av} = C_{pg} - C_{pl}$ (8)

2.1.2.2. Riser Model
The statement of the conservation of mass of species, i, is as follows:
Rate of accumulation of mass of species i = (Input rate – output rate + rate of production) of species i.
These terms may, thus, be expressed in terms of the process and spatial variables, and time, t, as:

Rate of accumulation = $\frac{\partial}{\partial t}\left(y_i\rho_c(1-\varepsilon_c)A_R\Delta L\right)$

Input rate = $F_{gR}y_i\big|_L$

Output rate = $F_{gR}y_i\big|_{L+\Delta L}$

Rate of production = $(-r_i)\rho_cA_R(1-\varepsilon_c)\Delta L$ ($-r_i$ being the rate, per unit volume, of gas oil cracking reactions producing species i.)
Thus,

$$\rho_c(1-\varepsilon_c)A_R\Delta L\frac{\partial yi}{\partial t}=F_{gR}y_i\big|_L-F_{gR}y_i\big|_{L+\Delta L}+(-r_i)\rho_c(1-\varepsilon_c)A_R\Delta L \quad (39)$$

But

$$y_i\big|_{L+\Delta L} = y_i\big|_L + \frac{\partial y_i}{\partial L}\Delta L \quad (10)$$

Substituting (10) in (9) and dividing through by the coefficients of $\frac{\partial y_i}{\partial L}$, the mass balance equation becomes

$$\frac{\partial y_i}{\partial t} = \frac{-F_{gR}}{\rho_c(1-\varepsilon_c)A_R}\frac{\partial y_i}{\partial L} + (-r_i) \quad (11)$$

In terms of dimensionless spatial variable, $z = \frac{L}{L_R}$,

$$\frac{\partial yi}{\partial t} = \frac{-F_{gR}}{\rho_c(1-\varepsilon_c)A_R L_R}\frac{\partial yi}{\partial z} + (-r_i)) \quad (12)$$

The volumetric flow rate of hydrocarbon in the riser, and that of the catalyst are determined from equations (13) and (14) respectively, as,

$$V_g = \frac{F_{gR}}{\rho_{gR}} \quad (13)$$

$$V_c = \frac{F_{rc}}{\rho_c} \quad (14)$$

From which the hydrocarbon volume fraction may be deduced as

$$\varepsilon_{gr} = \frac{V_g}{V_g + V_c} \quad (15)$$

$$\varepsilon_{gr} + \varepsilon_c = 1 \quad (16)$$

The density of the hydrocarbon phase is a function of the temperature in the riser and may be obtained, assuming ideal gas behaviour, from equation (3.16)

$$\rho_{gR} = \frac{P_{rrg} M_{gr}}{RT_{rx}} \quad (17)$$

2.1.2.3 Riser Energy Balance

Over the same control volume employed in the species mass balance.

accumulation of energy = Energy input rate – energy output rate + production rate of energy.

Accumulation = $\left(\rho_{gR}C_{pg}\varepsilon_{gr} + \rho_c C_{pc}(1-\varepsilon_{gr})\right)\frac{\partial T_{rx}}{\partial t}.A_R\Delta L$ 18)

Input rate = $(F_{gr}C_{pg} + F_{rc}C_{pc})T_{rx}\big|_L$ (19)

Output rate = $(F_{gr}C_{pg} + F_{rc}C_{pc})T_{rx}\big|_{L+\Delta L}$ (20)

Energy production = $\left(\sum(-r_i)\Delta H_R\right)(1-\varepsilon_c)\rho_c A_R \Delta L$ (21)

Following the same approach that was employed in the mass balance case, the rate of change of temperature becomes:

$$\frac{\partial T_{rx}}{\partial t} = \frac{-\left(F_{gr}C_{pg} + F_{rc}C_{pc}\right)}{\left(\rho_{gR}C_{pg}\varepsilon_{gr} + \rho_c C_{pc}(1-\varepsilon_{gr})\right)A_R L_R}\frac{\partial T_{rx}}{\partial z} + \frac{(1-\varepsilon_c)\rho_c\sum\Delta H_R(-r_i)}{\left(\rho_{gR}C_{pg}\varepsilon_{gr} + \rho_c C_{pc}(1-\varepsilon_{gr})\right)} \quad (22)$$

2.1.2.4 Separator Subsystem Model

The significance of the separator is to cause a time delay between the riser and the regenerator operations. However, coke concentration on the catalyst leaving the separator and the temperature in the separator are two variables of interest. A continuously stirred tank model is applied to describe the dynamics of the separator (stripper).

Coke Balance:

ACC = Input rate - output rate.

$$\frac{d(C_{st}\ M_{st})}{dt} = F_{rc}y_4 - F_{sc}C_{st}. \quad (23)$$

$$\frac{dC_{st}}{dt} = \frac{1}{M_{st}}\left[F_{rc}y_4 - F_{sc}C_{st}\right] \quad (24)$$

Energy Balance:

$$\frac{dH}{dt} = \dot{H}_{in} - \dot{H}_{unt}. \quad (25)$$

$$\frac{d\left(M_{st} C_{pc} T_{st}\right)}{dt} = F_{rc}\ C_{pc}\ T_{rx} - F_{sc}\ C_{pc}\ T_{st}. \quad (26)$$

$$\frac{dT_{st}}{dt} = \frac{1}{M_{st}}\left(F_{rc}\ T_{rx} - F_{sc}\ T_{st}\right) \quad (27)$$

2.1.2.5: Regenerator Models

The regenerator is modelled as a continuously stirred tank reactor in which the coke concentration and temperature are uniformly distributed. This follows the work Hovd and Skogestad (1993); Jens et al. (1992). Thus, the coke mass balance gives:

$$\frac{d\left(M_{rg} C_{rc}\right)}{dt} = F_{sc}\ C_{st} - F_{rc}\ C_{rc} - r_{cb} \quad (28)$$

$$\frac{dC_{rc}}{dt} = \frac{1}{M_{rg}}\left(F_{sc}\ C_{st} - F_{rc}\ C_{rc} - r_{cb}\right) \quad (29)$$

Oxygen in regenerator dense bed

$$\frac{d\left(M_{or} O_d\right)}{dt} = \frac{F_a}{M_a}\left(O_{in} - O_d\right) - \alpha r_{cb} \quad (30)$$

$$\frac{dO_d}{dt} = \frac{1}{M_{or}}\left(\frac{F_a}{M_a}\left(O_{in} - O_d\right) - \alpha r_{cb}\right) \quad (31)$$

Temperature in regenerator dense bed:

$$\frac{dT_{rg}}{dt} = \frac{1}{C_{pc} M_{rg}}\left(T_{st} F_{rc} C_{pc} + T_a F_a C_{pa} - T_{rg}\left(F_{rc} C_{pc} + F_a C_{pa}\right) - \frac{\Delta H r_{cb}}{MC}\right) \quad (32)$$

2.2 Model Solution and Validation

The riser partial differential equations were discretized in space using the method of lines with eighteen (18) internal nodes and two boundary nodes, giving eighty (80) ordinary differential equations, odes for species concentration and twenty (20) odes for riser temperature. The100 riser odes, along with the regenerator equations and other constitutive relations given in the model development section were then solved in MatLab for species concentration, riser temperature and regenerator temperature respectively. Design and operating data and results from previous studies were collected for model validation, methods validation, and scenario comparison. The collected data were extracted from monographs, empirical correlations, graphs, and tables and presented here in tabular format. Thermo-physical properties of species are given in table 1 while the basic dimensions of the FCC plant of reference refinery, studied in this paper are as shown in table 2. Kinetic parameters are presented in table 3 while case-specific operating data are presented in table 4

Table1: Thermo-physical properties of reacting species [(NPRC, (1987); Josiah et.al, (2003)]

Parameter	**Value**
Hydrocarbon Vapour Density, kg/m3	9.52
Liquid density at 288K, kg/m3	924.8
Specific heat of Hydrocarbon vapour, kJ/kg K	3.3
Specific heat of feed, kJ/kg K	2.67
Heat of vaporization, kJ/kg	156

Vaporization temperature, K	698
Catalyst bulk density, kg/m3	925
Particle size, μm	7.5
Catalyst heat capacity, kJ/kg K	1.12

Table 2 : Dimensions of Industrial FCC Riser-Regenerator [NPRC. (1987)]

Parameter	**Value**
Riser Length, m	22.9
Riser Diameter, m	2.9
Regenerator length, m	35.45
Regenerator diameter, m	9.8
Cyclone Height, m	14.24
Cyclone diameter, m	1.5
Disengagement height, m	24.5

Table 3: Base Kinetic Parameters for Four-Lump reaction scheme**. [**Josiah and Etebu (2006)**]**

Reaction	**Pre-exponential factor, s**	**Activation Energy (kJ/Kmol)**
Conversion of gas oil to gasoline	221.611	68249.6
Conversion gas oil to light gases	1263.611	89216.4
Conversion of gas oil to coke	10.4583	64575
Conversion of gasoline to light gases	0.90417	50718.4
Conversion of gasoline to coke	2210.2778	115458
Catalyst decay	83806.556	117705

Table 4 Data used for Case Comparison [Ali et.al (1997; Mehran et al (2010)]

Variable	Case 1	Case 2	Case 3	Case 4
Riser Length, m	33	33	33	33
Riser Diameter, m	0.8	0.8	0.8	0.8
Regenerator Length, m	11	11	11	11
Regenerator Diameter, m	5.8	5.8	5.8	5.8
Feed quality, API^o	21.76	22.98	22.73	22.28
Feed Rate, kg/s	25.7	26,93	23.61	19.95
Catalyst/Oil(kg/kg)	6.33	5.43	6.07	7.24
Air Rate(kg/s)	15.8	15.8	15.8	15.8
Air inlet Temperature(K)	378	378	378	378
Feed Temperature(K)	494	494	494	494

Data shown in Table 1, Table2, Table 3 and Table 4 were implemented as MatLab m-files and used in the algorithm for model solution. A dynamic solvers and simulator (**twentypointsdynamics.m**)was developed for the solution of the modelling equations in a sequential-modular framework. The simulator consists of a main function file, six sub-functions and a script file in which the script acts as a container that houses the main function while the main function houses the sub-functions. The computational structure that was employed in the solver-simulator is as follows

Open Script file
Call main function
Call sizes input file
Call operating conditions file
Call constitutive relations file
Call Modelling equations file
Invoke ode23 solver
Call results generator file
Close script file

3: Results and Discussion

3.1: Comparison with industrial FCC unit data

A comparison of model predictions with steady state plant data is given in Table 5. According to this study, the model predictions for gasoline yield, coke yield, riser temperature and regenerator temperature are minimally higher than observed plant data. On the other hand, model predictions relative to observed data are minimally lower for coke on regenerated catalyst, flue gas oxygen concentration, light gases yield and concentration of unconverted gas oil feed. The deviations of the predictions from plant data are well below 15%, hence not significant. On this premise, we deduce that model predictions show good agreement with plant data from the base case refinery. It has been established elsewhere that a high regenerator temperature would lead to a high riser temperature, high gasoline yield, high gas oil conversion low flue gas oxygen concentration and low coke concentration on regenerated catalyst The model predictions , relative to the plant data, follow the established FCC unit behaviour that was alluded to in Ansari (2010),Kuma(2014). The observed trend can be explained in the context of coke burn reactions in the regenerator. Coke burn reactions in the complete combustion mode employ coke and oxygen in the regenerator, causing temperature rise due to the release of heat from the reactions. The implication is that there would be less coke on the regenerated catalyst entering the riser as well as less oxygen in the flue gas leaving the regenerator.

Table 5: Comparison of Model Predictions with Steady State Plant Data

Variable	Plant Data	This study	Error	% Error
Gasoline yield (wt %)	45.9	47.74	-1.84	-4.008
Light Gases yield (wt %)	26.6	25.65	0.95	3.571
Coke yield (wt %)	5.1	5.38	-0.28	-5.490
Unconverted Feed (wt %)	22.4	21.56	0.84	3.75
Riser Outlet Temperature (^{o}C)	524	527	-0.03	-0.573
Regenerator Temperature (^{o}C)	743	774	-31	-4.172
Coke on regenerated catalyst (wt %)	0.05	0.046	0.004	8.
Oxygen in flue gas (mole %)	3.0	2.84	0.16	5.33

3.2: Comparison with Previous work

The results from this work in comparison with those from literature are as given in table 6,table 7, table 8 and table 9 respectively. According to the analysis in table 6, model predictions in respect of riser temperature and regenerator are better, compared to literature, More so, the deviation from plant data favours the results from this work especially for gasoline yield, coke yield and oxygen concentration. In a similar vein, temperature predictions from this study are more reliable in comparison with data from literature, as shown in table 7, table 8 and table 9 respectively

Table 6: Analysis of Results for Model Validation Case 1

Vessel/Measure	Value		Deviation from Plant Data		Relative direction of Prediction	
	Literature	This study	Literature	This study	Literature	This study
Gasoline yield (wt %)	48.01	47.63	1.41	1.03	Higher	Higher
Coke yield (wt %)	5.41	5.27	0.07	-0.07	Higher	Lower
Riser Temperature (K)	845	812	37	4	Higher	Higher
Oxygen concentration (mole %)	2.60	4.17	0.3	1.87	Higher	Higher
Regenerator Temperature (K)	1121	1011	88	-22	Higher	Lower

Table 7: Analysis of Results for Model Validation Case 2

Vessel/Measure	Value		Deviation from Plant Data		Relative Direction of Prediction	
	Literature	This study	Literature	This study	Literature	This study
Gasoline yield (wt %)	45.96	47.60	3.17	4.84	Higher	Higher
Coke yield (wt %)	5.11	5.26	-0.32	-0.17	Lower	Lower
Riser Temperature (K)	826	809	21	4	Higher	Higher
Oxygen concentration (mole %)	3.53	4.1	0.5	0.9	Higher	Higher
Regenerator Temperature (K)	1117	1003	113	-1	Higher	Lower

Table 8: Analysis of Results for Model Validation Case 3

Vessel/Measure	Value		Deviation from Plant Data		Relative direction of Prediction	
	Literature	This study	Literature	This study	Literature	This study
Gasoline yield (wt %)	48.51	46.49	6.73	4.71	Higher	Higher
Coke yield (wt %)	5.47	5.00	-0.22	-0.69	Lower	Lower
Riser Temperature (K)	835	801	29	-5	Higher	Lower
Oxygen concentration (mole %)	3.13	3.01	0.23	0.11	Higher	Higher
Regenerator Temperature (K)	1117	1026	108	17	Higher	Higher

Table 9: Analysis of Results for Model Validation Case 4

Vessel/Measure	Value		Deviation from Plant Data		Relative direction of Prediction	
	Literature	This study	Literature	This study	Literature	This study
Gasoline yield (wt %)	51.30	46.48	7.47	2.65	Higher	Higher
Coke yield (wt %)	5.79	5.81	-0.004	-0.02	Lower	Lower
Riser Temperature (K)	843	799	50	4	Higher	Higher
Oxygen concentration (mole %)	2.75	3.1	-0.25	0.1	Lower	Higher
Regenerator Temperature (K)	1143	964	183	4	Higher	Higher

3.3: Steady State Concentration and Temperature Profiles

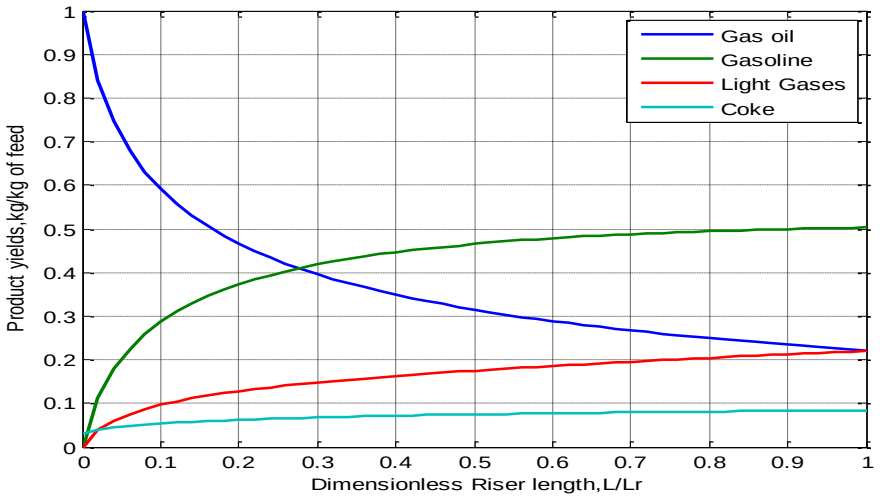


Figure 3: Axial Concentration Profiles in the Riser

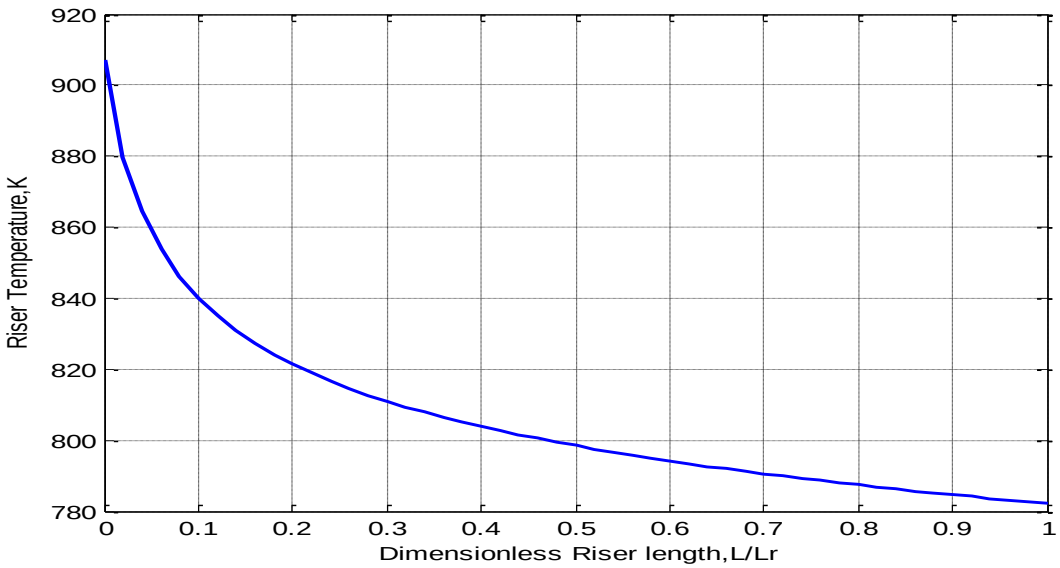


Figure 4: Axial Temperature Profile in the Riser

A qualitative analysis in which the predicted responses are compared based on trends and trajectories has been employed in this paper, as consistent with the work of Chang (2001), Gupta (2011) and Prabha (2014). Fig 3 shows the yields of the key products and un-converted gas oil along the length of the riser at steady state. The overall conversion of gas oil in the riser is 77.81%, of which 62.68% of the initial feed was converted within 0 to 4.58 meters of the riser length. This length of the riser represents the first 20% of the entire length of the riser. In the same manner, gasoline yield rose rapidly from zero at the mixing zone to about 40% within 4.58 meters away from the mixing zone, representing 80% of the total yield (47%). The axial variation of temperature at steady state is as shown in Fig 4.2. Riser temperature dropped exponentially from a mixing point value of 907 K (632^{o}C) to 782 K, representing 125^{o}C drop. It is of interest to note again, that about 70% of the temperature drop occurred within the first 20% of the riser length. The observed steady state behaviour of the FCCU as reported in this work is in tandem with literature (Gupta, 2011). Mehran (2010), Fernandes et al (2007), (2008), Prabha 2012). More so, it was observed that model predictions for gasoline

yield, riser temperature and oxygen concentration are consistently higher than plant data while predictions for coke yield and regenerator temperature did not follow a trend. The observed deviations from plant data were principally due to paucity of data, as a generic thermo-physical data set was used for the different case scenarios in the validation step

4: CONCLUSION

The best approach to conducting simulation study is to engage an industrial unit or a pilot scale unit. However, due to safety, cost, operational and other constraints, these options are not readily available, hence the search for suitable models. In this regard, a control-relevant dynamic model of the fluid catalytic cracking unit has been developed and validated in this paper using data from an industrial FCC unit and literature. In the light of the results obtained and the observed process behaviour, this paper concludes as follows:

1. Model predictions are in very close agreement with industrial FCC unit data
2 The observed FCC unit behaviour is in tandem with literature
3 The developed and validated model is suitable for evaluating control configurations and testing control algorithms.
4. Over 62 percent of gasoil conversion is achieved in the first 20 percent of the riser height.
5 At least 70 percent of temperature drop occurs in the first 20 percent of the riser height.
6 The first 20 percent of the riser height is the region that is most important for evaluating heat and mass transfer characteristics.

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