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Steady State Simulation of a Fluid Catalytic Cracking Unit

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Abstract: In the present study a steady state simulation for the fluid catalytic cracking was investigated. A mathematical model for the reactor and regenerator was developed and they were coupled using enthalpy balance between the reactor and the regenerator. The reactor was modelled as a plug flow reactor incorporating the 4-lump model for cracking reactions whereas the regenerator was modelled as a fluidised incorporating the coke burning kinetics. For the regenerator, the reaction taking place in the free board was also taken into consideration. The developed model has been validated using the operating data of an industrial FCCU reported in the literature. Some of the model parameters have been estimated. The predicted data coincide reasonably with the experimental data.

Key words: Simulation, fluid catalytic cracking, kinetics of coke burning, dense bed model, free bed model

INTRODUCTION

Fluid Catalytic Cracking Unit (FCCU) is one of the most important processes in the petroleum refinery, being the most frequently employed operation for the conversion of gas oil, atmospheric and/or vacuum residues into more useful high octane gasoline.

The FCCU is a quite complex unit with its performance linked to many independent and dependent operating variables. Several models were reported in the literature (Ali and Rohani, 1997; Aranoldes and De Lasa, 1992; Arthur, 1951). Kumar *et al.* (1995) developed a comprehensive integrated simulator to simulate the riser and the regenerator. The reactor modelled as a plug flow reactor by incorporating the 10-lump kinetic model of Jacob *et al.* (1976). Similar models were developed by incorporating the 4-lump kinetic model of Lee *et al.* (1989). The regenerator models were developed by investigators (Morley and De Lasa, 1987; De Lasa, 1981; Guigen and Large, 1984; Krishna and Parkin, 1985). A mathematical model for the heavy oil cracker regenerator bed was reported (Wells *et al.*, 1992). Wei *et al.* (1993) have studied the mixing of gas, solid and heat in commercial FCC regenerator. A one dimensional pseudo-dispersion model for the regenerator was developed and the model predictions were compared with the axial temperature and concentration profiles in a commercial unit. This study indicates that mixing of heat and mass is quite significant

in a large fluidised regenerator. Simple models based on two-phase theory have been used by Elnashaie and El-Hennawi (1979) in the analysis of multiple steady states for fluidised bed catalytic cracking units.

Operating limits selected for these variables depend upon the unit as well as its operating policy and thereby add to the complexities of the control. The objective of the FCCU operation is to maximise the yield of high octane gasoline and minimise coke formation to make it economically attractive. Further, establishment of the optimum conditions by performing trial and error experiments on a running operating conditions by process simulation which eliminates the inherent risks associated with the actual experimentation on a running plant.

Hence, in the present study a steady state simulation model is developed for the FCCU. This is accomplished by developing mathematical models for the reactor and the regenerator of the FCCU and then interconnecting the. The reactor is modelled as a plug flow reactor incorporating the four lump kinetic model for the catalytic cracking reactions and the regenerator as fluidised bed incorporating the coke burning kinetics. The models for the reactor and the regenerator are then interconnected by developing enthalpy balance equations between the reactor and the regenerator to form the simulation model for the FCCU. This simulation model is then validated using the operating data of a commercial scale FCCU reported in the literature.

MATHEMATICAL MODEL

There are several models reported in the literature for the kinetic scheme of the cracking reactions. In this work, the 4-lump model by Lee *et al.* (1989) has been selected because of its ability to calculate the coke formation rate independently. The model assumes that the gas oil is cracked into the most desired gasoline and the by-products C₁-C₄ gas and coke. Since the riser reactor is operating at high temperatures, the secondary cracking occurs for gasoline to form coke and C₁-C₄ or Light Hydrocarbon Gas (LHG). The assumptions involved are: gas flow in the reactor is in ideal plug flow, axial dispersion in the reactor is negligible, gas oil cracking is second order reaction, gasoline cracking is first order reaction, both gas oil and gasoline have identical activity decay function, C₁-C₄ gas does not produce coke, which is mainly produced from aromatic hydrocarbons, coke content in feed is very low and reactions are almost completed in riser.

Mass balance:

Gas oil:

$$\frac{dy_o}{dz} + \frac{\Phi_r A_r L_r \epsilon_r \rho_g}{F_g} [(K_{ob} + K_{oc} + K_{od})y_o] = 0 \quad (1)$$

Gasoline:

$$\frac{dy_g}{dz} + \frac{\Phi_r A_r L_r \epsilon_r \rho_g}{F_g} [(K_{gc} + K_{gd})y_g - (K_{ob} \times y_o^2)] = 0 \quad (2)$$

Light Hydrocarbon Gases:

$$\frac{dy_d}{dz} - \frac{\Phi_r A_r L_r \epsilon_r \rho_g}{F_g} [K_{gd} \times y_g + K_{oc} \times y_o^2] \quad (3)$$

Coke:

$$\frac{dy_c}{dz} - \frac{\Phi_r A_r L_r \epsilon_r \rho_g}{F_g} [K_{bc} \times y_b + K_{oc} \times y_o^2] \quad (4)$$

Energy balance:

$$\frac{dT_r}{dz} + \frac{\Phi_r A_r L_r \epsilon_r \rho_g}{(F_g \times C_{pg} + F_c \times C_{pc})} \left\{ \begin{aligned} & y_o^2 [K_{ob} \times H_{nb} + K_{od} \times H_{nd} + K_{oc} \times H_{rc}] + \\ & y_g [K_{bc} \times H_{rc} + K_{bd} \times H_{rd}] \end{aligned} \right\} = 0 \quad (5)$$

with the boundary conditions

$$\left. \begin{aligned} y_i(0) &= 1 \\ y_i(0) &= 0 \text{ for } i = b, c, d \\ T_r(0) &= \text{Initial temperature after mixing} \end{aligned} \right\} \quad (6)$$

The above ordinary differential equations are solved simultaneously by the fourth order Runge-Kutta method using the kinetic and thermodynamic parameters as given by Ali and Rohani (1997).

Regenerator model: The deactivated catalyst from the reactor passes through the spent catalyst slide valve and then enters the regenerator at the bottom. Catalyst flow into regenerator may be tangential as well as downward. Major purpose of regenerator is to oxidize the coke on the spent catalyst with oxygen to form CO, CO₂ and H₂O thereby regenerating the catalyst. After compression, combustion air enters at the bottom of regenerator through a grid distribution pattern designed to provide efficient mixing of air with deactivated catalyst and this results in a fluidized bed. Regenerated catalyst leaves through a standpipe which is circumferentially remote from spent catalyst inlet in order to avoid short circuiting. The nearly coke free catalyst, after passing through the regenerated slide valve is mixed with oil at riser base and the cycle is repeated. Provisions are made for adding fresh catalyst makeup to maintain inventory and for withdrawal of aged and contaminated catalyst.

The regenerator consists of two zones: a high density fluidized bed which is known as the dense bed and a dilute phase zone known as freeboard region. The dense bed is again considered to have two phases, the emulsion phase which has very high catalyst density and the bubble phase, which has low catalyst density. Gaseous combustion products and some entrained catalyst are conveyed out of the regenerator dense bed into a dilute phase zone known as free board region, which sets of primary and secondary cyclones remove and return the catalyst to the dense bed. A schematic diagram of the regenerator is shown in Fig. 1.

The overall mathematical model of the regenerator consists of two sub models, viz., the dense bed model and the freeboard model. We apply the two region model for the dense bed which consists of bubble phase and emulsion phase. Above the dense bed region is the freeboard region characterized by low catalyst entrainment.

The following assumptions are made in the development of the regenerator model:

Air entering into the regenerator splits into two parts; one passed through emulsion phase to keep the catalyst

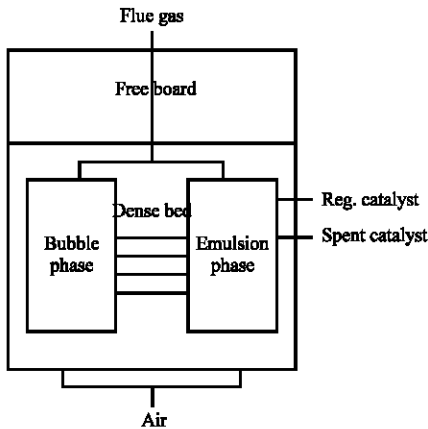


Fig. 1: Schematic diagram of a regenerator

particles at minimum fluidization condition and the other part which is in excess of this minimum fluidization condition passes through the dense bed as bubbles.

The bubble phase and the freeboard are in plug flow whereas the emulsion phase is fully mixed. All entrained catalyst returns to dense bed through cyclones. Bubble phase does not contain any solids and no chemical reactions take place in bubble phase. All chemical reaction occurs in the emulsion phase and freeboard. Ideal gas is applicable. Regenerator is adiabatic.

Kinetics of coke burning: The coke which is a mixture of carbon and hydrogen undergoes the following chemical reactions in the emulsion phase and freeboard.



Reactions (1) and (2) are intrinsic carbon combustion reaction on catalyst. Both reactions follow second order kinetics. Most kinetic data are reported for the overall carbon consumption i.e. sum of reactions (1) and (2). In our present study the kinetic expression given by Morley and De Lasa (1987) was selected for overall carbon consumption. The CO/CO₂ ratio is calculated using semi-empirical correlation given by Errazu *et al.* (1979).

Dense bed model: As in the case of the reactor, the methodology adopted for the regenerator is also to write the material and energy balance for a differential section of the regenerator and then integrate the differential equations using proper initial conditions. For easy representation the components are numbered in the following order:

$$1 = CO, 2 = CO_2, 3 = H_2O, 4 = O_2, 5 = N_2$$

The dense bed consists of bubble phase and emulsion phase and the bubble phase is modeled as plug flow reactor and the emulsion phase as a CSTR.

As the amount of catalyst present in the bubble phase is negligible, no gas-solid chemical reaction takes place in the bubble phase. Hence, the change in concentration of the various components is due to mass transfer between bubble phase and emulsion phase.

The following material and energy balances can be written for a differential height, dx, for the bubble phase.

Material balance: Material balance of CO, CO₂, H₂O and O₂ are obtained from the following equations:

$$\frac{df_{b,i}}{dx} = -H_{ib} \times K_{m,i} \left[\frac{f_{b,i}}{U_{ib}} - \frac{f_{e,i}}{U_{e,i}} \right] \quad I = 1, 2, 3, 4 \quad (10)$$

where the bubble to gas interchange coefficient is obtained from the following expression:

$$K_{m,i} = \frac{\delta}{\frac{1}{K_{e,i}} + \frac{1}{K_{c,i}}} \quad (11)$$

The bubble fraction of the dense bed, δ is obtained from the equation:

$$\delta = \frac{U_{gbo}}{U_{go} - U_t + U_{br} + U_f} \quad (12)$$

The bubble to cloud and cloud to emulsion gas interchange coefficient for component i are calculated using the following expression.

$$K_{bc,i} = \frac{4.5U_{ci}}{D_b} + \frac{5.85D_b^{0.5}g^{0.25}}{D_b^{1.25}} \quad (13)$$

$$K_{ce,i} = 6.78 \left[\frac{\epsilon_e D_e U_b}{D_b^{1.25}} \right]^{0.5} \quad (14)$$

Energy balance: The temperature of the bubble phase is calculated from,

$$\frac{dT_b}{dx} = - \frac{H_i \times A_{ib} \times H_{ib} (T_b - T_i)}{\sum_{i=1}^5 f_{b,i} C_{p,i}} \quad (15)$$

The bubble to emulsion heat interchange coefficient is calculated according to the following equation:

$$H_i = \frac{4.5U_e \rho_e C_{pe}}{D_b} + \frac{5.85(K_e \rho_e C_{pe})^{0.5} g^{0.25}}{D_b^{1.25}} \quad (16)$$

The initial conditions for the integration of the above equation are given by:

$$\left. \begin{aligned} F_{b,i} &= 0, i = 1, 2, 3 \\ F_{b,s} &= 0.21Fab \\ T_b &= T_i \\ U_{bo} &= U_{bo} \end{aligned} \right\} \quad (17)$$

The various hydrodynamic parameters required for the solution of the model equations are evaluated as follows. The various velocities are obtained by the following equations given by Kunii and Levenspiel (1990).

$$U_{so} = \frac{FM_s}{\rho_s A_{so}} \quad (18)$$

$$U_s = \frac{F_s}{\rho_s A_{so}} \quad (19)$$

$$U_{br} = 0.711(gD_b)^{0.5} \quad (20)$$

$$U_{soo} = U_s + U_{br} \quad (21)$$

$$U_{bo} = U_{so} + U_{soo} \quad (22)$$

$$\frac{dU_{bo}}{dx} = \frac{R \times H_{ab} \times H_i (T_b - T_s)}{P \times C_{pm}} \quad (23)$$

With the assumption that all the entrained catalyst returns to dense bed through cyclone, an overall material balance for the emulsion phase can be written and carbon concentration in the emulsion phase can be calculated as

$$C_s = \frac{F_s C_o}{W_c K_{s2} C_{so} + F_s + E} \quad (24)$$

Similarly hydrogen concentration in the emulsion phase can be calculated as,

$$H_s = \frac{F_s H_o + E H_i}{W_c K_s C_{so} + F_s + E} \quad (25)$$

An overall energy balance can be written for dense bed and from which temperature of emulsion phase can be obtained as,

$$T_i = \frac{\left\{ T_{inf} + 28F_s C_{ps} (T_s - T_{inf}) + EC_{ps} (T_f - T_{inf}) + \right.}{F_s C_{ps} + EC_{ps}} \left. \left\{ F_s C_{ps} (T_c - T_{inf}) - F_s C_{ps} (T_b - T_{inf}) + H_{ts} \right\} \right\} \quad (26)$$

The flow rates of each component in the emulsion phase are then estimated using the following equations:

$$F_{s1} = \frac{C_{so} F_s (1 - H_{oc}) X_c K_2}{12(1 + K_2)} \quad (27)$$

$$F_{s2} = \frac{C_{so} F_s (1 - H_{oc}) X_c}{12(1 + K_2)} \quad (28)$$

$$F_{s3} = \frac{C_{so} F_s (H_{oc}) X_c}{2} \quad (29)$$

where

$$X_c = 1 - \frac{C_s}{C_o} \quad (30)$$

$$X_h = 1 - \frac{H_s}{H_o} \quad (31)$$

Freeboard model: Bubbles erupt at the surface of the dense fluidized bed, ejecting catalyst particles into freeboard region. The product gases from the bubble and emulsion phase of the dense bed along with entrained catalyst thus enter the freeboard region. The overall reaction rate in the large scale fluidized bed depends not only on what goes on in bed, but also on reaction occurring in freeboard region. The influence of the freeboard is due to the presence of unreacted carbon and hydrogen in entrained catalyst and oxygen in flue gas. The following material and energy balance equations can be written for a differential height, dx, of the freeboard. For $i = 1, 2, 3, 4$,

$$\frac{df_{i,s}}{dx} = A_b H_b (1 - \epsilon_i) r_i \quad (32)$$

$$r_1 = \frac{C_f C_{f,Ab1} + 1}{1 + K_1} \quad (33)$$

$$r_2 = \frac{C_f C_{f,Ab1} + 1}{1 + K_1} \quad (34)$$

$$r_3 = K_1 H_i C_{f,s} \quad (35)$$

$$C_{f,s} = \frac{f_{f,s}}{A_b U_g} \quad (36)$$

$$\frac{df_{f,s}}{dx} = A_b H_b (1 - \epsilon_i) \left[r_1 + \frac{r_1 + r_2}{2} \right] \quad (37)$$

$$\frac{dC_f}{dx} = \frac{A_b H_b (1 - \epsilon_i) (r_1 + r_2)}{E} \quad (38)$$

$$\frac{dT_r}{dx} = \frac{A_b H_b (1 - \varepsilon_r) \sum r_i (-H_i)}{EC_{p_r} + \sum f_{i,d} C_{p,i}} \quad (39)$$

The above equations are solved by the fourth order Runge-Kutta method and the initial values for the above equations are obtained after solution of the dense bed equations.

Computational procedure for regenerator model: The solution of the regeneration model involves iterative calculations and the step by step procedure is described below:

Read input data

Assume T_b the flue gas outlet temperature.

Assume $C_p H_f$ the carbon and hydrogen concentration in the entrained catalyst that leaves the freeboard and returns to the dense bed through cyclones.

Assume T_e the emulsion phase temperature.

Assume $C_{e,4}$ the oxygen concentration in the emulsion phase. Also calculate the amount of CO , CO_2 , H_2O formed.

Integrate the oxygen mass balance equation for the bubble phase by fourth order Runge-Kutta method and check whether the oxygen mass balance satisfies the pre-assigned tolerance or not. If not satisfies proceed to step 8. Otherwise update assumed number of moles of oxygen in emulsion phase by interval having method and repeat from step 6 to till convergence is obtained.

Assume the moles of CO , H_2O present in the emulsion phase and calculate moles of CO_2 and integrate the differential mass balance equations for CO , H_2O , CO_2 for bubble phase and check whether the mass balance equation for carbon and hydrogen satisfy with in a pre-assigned tolerance or not. If satisfies proceed to step 9, else update assumed number of moles of CO and H_2O in emulsion phase, by interval halving method and repeat calculation till convergence is obtained.

Calculate the new emulsion phase temperature and check whether the new emulsion phase temperature and check whether the new emulsion phase temperature matches with the assumed value or not. If it matches, proceed to step 10, otherwise repeat calculations form step 4 onwards till convergence is obtained.

Calculate entrainment rate and integrate the freeboard equations and check whether the carbon and hydrogen concentrations in the entrained catalyst matches with the assumed value of not. If they match, proceed to step 11, otherwise repeat calculations from step 4 onwards till convergence is obtained. Check whether flue gas outlet temperature obtained by solving freeboard equations

match with the assumed value or not. If it matches proceed to step 12, otherwise repeat from step 5 onwards till convergence is obtained.

MODEL VALIDATION AND SIMULATION

Model validation: The first step in any simulation study is to check the accuracy of the model developed by comparing its predictions with the operating data of a commercial plant and adjust some of the model parameters such as the model predictions match reasonably well with the plant data. During the course of present study we could not collect the operating data of any commercial FCCU. Hence in the present work model validation study is undertaken by comparing the model prediction with the operating data of a commercial FCCU reported in the open literature. Table 1 and 2 show, respectively, the dimensions of the FCCU, operating data and physical properties and parameters used in the model validation Ali *et al.* (1967). Table 3 shows a comparison of the model predictions with one set of operating data of a commercial FCCU reported by Ali *et al.* (1967) and it is clear from the Table 3 that the agreement obtained between the model predictions and plant data is quite good.

The model parameters selected in this study are the multiplication factors for the frequency factors of the rate equations. This was necessary because the kinetic parameters as used from the literature Lee *et al.* (1989) depend on the catalyst type, the catalyst activity, the age of the catalyst in the operation and quality of the feed

Table 1: Dimensions of the fluid catalytic cracking unit

Height of the riser	33.0 m
Diameter of the riser	0.8 m
Height of the regenerator	11.0 m
Diameter of the regenerator	5.8 m

Table 2: Operating data of the fluid catalytic cracking unit

Gas oil flow rate	19.95 kg sec ⁻¹
Catalyst circulation rate	144.43 kg sec ⁻¹
Gas oil inlet temperature	221°C
Air flow rate	0.55 kg moles sec ⁻¹
Air inlet temperature	105°C
Regenerator pressure	2.20 kg/cm ² g
Coke H/C (wt. %)	12.40

Table 3: Results of model validation.

Riser	Plant	Calculated
Outlet temperature (°C)	522	574
Outlet composition (%)		
Gas oil	---	19.61
Gasoline	43.88	52.06
Light gas	---	22.21
Coke	5.83	6.12
Coke on deactivated catalyst (kg coke kg ⁻¹ catalyst)	---	0.00085
Regenerator		
Catalyst outlet temperature (°C)	687	675
Coke on deactivated catalyst (kg coke kg ⁻¹ catalyst)	---	0.00027

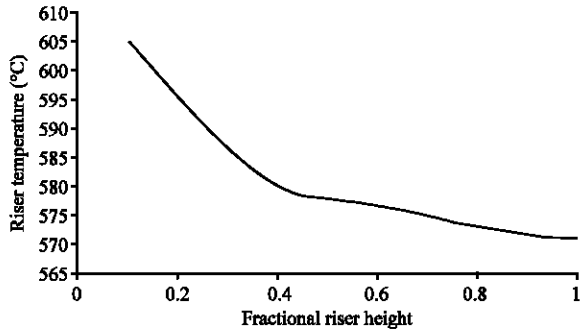


Fig. 2: Temperature profiles in the riser

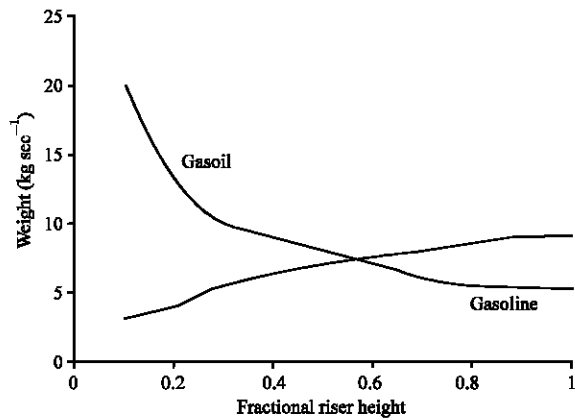


Fig. 3: Composition profiles in the riser

stock etc. Hence, the direct use of the kinetic parameters may not produce a good match between plant measurements and model predictions.

The temperature and composition of the gas mixture along the axial length of the riser are shown in Fig. 2 and 3. Since the cracking reactions are endothermic and the reactor is adiabatic temperature of the gas mixture drops as it moves up the reactor. The temperature of the regenerated catalyst entering the reactor is 691°C and mixing of the gas oil and catalyst results in the vaporization of the gas oil and as a result the temperature of the catalyst/oil mixture drops to 605°C and this is shown as the temperature of the gas mixture at the inlet of the reactor in Fig. 2. Composition of the gas oil, gasoline, light gases and coke are shown in Fig. 3. As expected composition of the reactant gas oil decreases and that of the product increases as the gas mixture passes upwards in the riser.

Simulation: In order to investigate the effect of the operating variables on FCCU performance a simulation study has been done. The effect of catalyst to oil ratio, air flow rate and gas oil inlet temperature on the FCCU

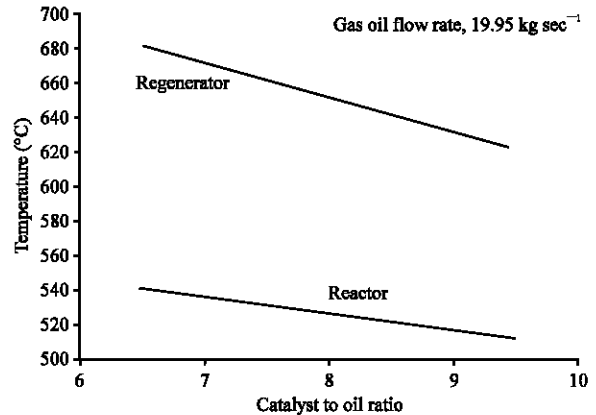


Fig. 4: Effect of catalyst to oil ratio on reactor and regenerator outlet temperature

performance has been investigated. The strategy adopted is to change only one variable at a time by keeping all the other variables at the base case values. Here, the base case values reports to values used in the model validation.

Catalyst to oil ratio: Catalyst to oil ratio can be varied either by changing the oil flow rate or catalyst circulation rate and in the present study both the options have been studied. Plant was operating with a catalyst circulation rate of 144 kg sec⁻¹ and oil flow rate of 19.95 kg sec⁻¹ so that the catalyst to oil ratio is 7.22. In the first case oil flow rate is maintained at 19.95 kg sec⁻¹ and catalyst circulation rate changed from 129.7 to 189.7 kg sec⁻¹ and results of the simulation are shown in Fig. 4. The reduced coke burning rate together with higher catalyst circulation rate results in the lowering of the regenerator temperature and consequently the reactor temperature also reduces as shown in Fig. 4. It can also be observed from Fig. 4 that the temperature difference between the regenerator and the reactor also decreases as the catalyst circulation rate increases.

Air flow rate: The air flow rate into the regenerator in the base case was 0.55 kg modes sec⁻¹. Simulation runs are made for various values of air flow rate and results are presented in Fig. 5 and 6. As the air flow rate increases temperature of the regenerator increases initially and then remain more or less constant and the reactor temperature increases only marginally. This marginal increase in reactor temperature does not affect the coke formation rate since all other operating parameters are same. Since the catalyst circulation rate is constant and the coke formation rate is same the coke content of the deactivated catalyst entering the regenerator is same for various values of air flow rate. As a result as the air flow rate

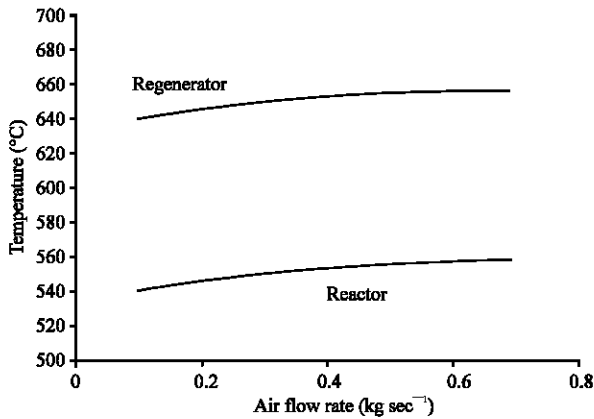


Fig. 5: Effect of air flow rate on reactor and regenerator temperature

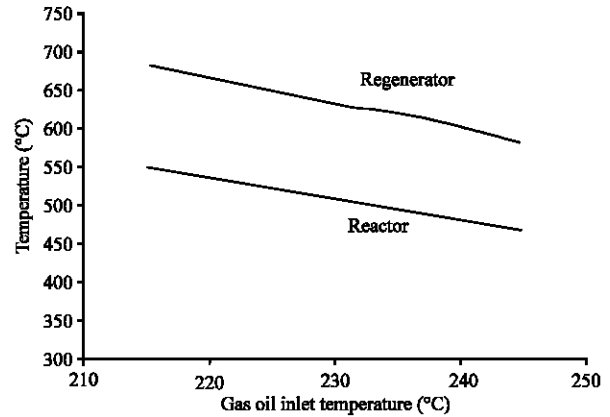


Fig. 7: Effect of gas oil inlet temperature on reactor and regenerator outlet temperature

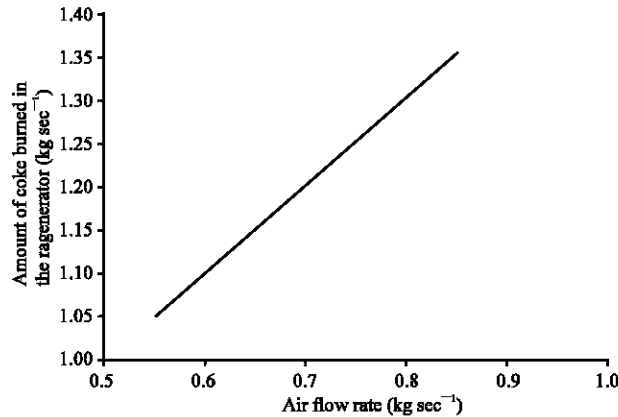


Fig. 6: Effect of air flow rate on coke burned in the regenerator

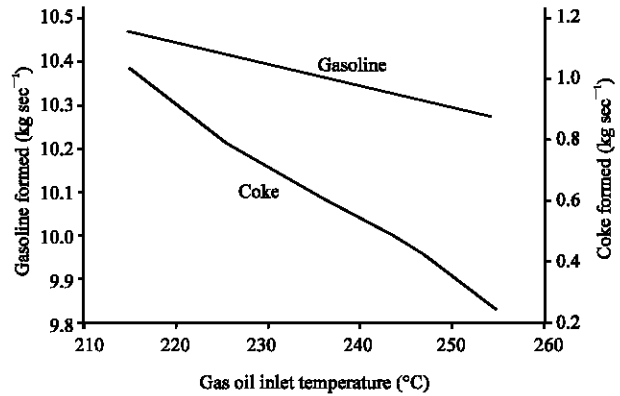


Fig. 8: Effect of gas oil inlet temperature on gasoline and coke formed

increases amount of coke burned in the regenerator increases until the complete coke is burned Fig. 6. This higher coke burning rate increases the temperature of the catalyst in the regenerator. It may be mentioned that once the coke conversion reaches the maximum value, further increase in air flow rate will cause the outlet temperature of the catalyst and flue gas to decrease since the amount of heat liberated by coke burning remains the same and more heat will be carried away by flue gas.

Gas oil inlet temperature: The inlet temperature of the gas oil into the reactor of the FCCU was 221°C as given in the operating data. In order to investigate the effect of change of the gas oil inlet temperature into the reactor a simulation study has been done. Simulations runs are made for various gas oil inlet temperatures into the reactor and the results of the study are shown in Fig. 7 and 8. It can be observed from Fig. 8 that as the gas oil inlet temperature into the reactor increases the amount of

gasoline and coke formed in the reactor decreases. Lower coke formation rate in the regenerator and this results in the temperature of the regenerator to decrease as the catalyst circulation rate is constant. This decreasing of the regenerator temperature causes the temperature of the reactor also to decrease as shown in Fig. 7.

CONCLUSIONS

In order to investigate the effect of change of some of the operating parameters on the FCCU performance, a simulation study has been carried out. The effect of change of the variables, catalyst to oil ratio, air flow rate and oil inlet on the performance has been studied.

Catalyst to oil ratio has been changed in two ways viz., changing the Catalyst Circulation Rate (CCR) by fixing the oil flow rate by fixing the CCR. In either case, it is observed that as the catalyst to oil ratio increases, the temperature of both reactor and regenerator decreases.

Decrease in temperature of the reactor results in decreasing the amount of gasoline and coke formed. As a result, coke burning rate in the regenerator decreases which in turn reduces the temperature of the catalyst leaving the regenerator.

It was observed that as the airflow rate increases the amount of oxygen available for coke burning increases and as a result the temperature of the regenerator increases initially and then slightly because of nitrogen build-up. Reactor temperature also increases slightly since the temperature of the regenerated catalyst increases.

As the gas oil inlet temperature increases the amount of gasoline and coke formed in the reactor decreases. This resulted in the coke burning rate in the regenerator to decrease which in turn reduces the regenerator temperature. Reduction in the regenerator temperature reduces the reactor temperature.

NOMENCLATURE

a Cyclone inlet height, m
 A_{db} Area of dense bed, m^2
 A_{fb} Area of freeboard, m^2
 b Cyclone inlet width, m
 C_{co} Coke content of catalyst, g.coke g^{-1} .catalyst.
 C_e Carbon content in the emulsion phase, kg sec^{-1} .
 C_{eo} Oxygen concentration in emulsion phase, kmol sec^{-1}
 C_f Carbon content in the entrained catalyst
 C_o Coke content of spent catalyst, g.coke g^{-1} . catalyst
 C_{pa} Specific heat of air
 C_{pg} Specific heat of gas
 C_{pgr} Heat capacity of gases in the riser
 C_{pi} Specific heat of component I, kcal/kmol K
 C_{pm} Specific heat of mixture
 C_{ps} Specific heat of catalyst, kcal/kg K
 D_b Bubble diameter, m
 dc Cone diameter at natural length
 D_{ib} Diffusivity of component I in bubble phase, $m sec^{-2}$
 d_{pi} Particle diameter
 E_a Activation energy, kcal mol^{-1}
 E Entrainment rate
 F_a Flow rate of air, kmol sec^{-1}
 F_{ab} Flow rate of air through bubble phase
 F_{ac} Flow rate of air through emulsion phase
 $F_{e,i}$ Flow rate of component i through emulsion phase, kmol sec^{-1}
 $F_{f,i}$ Flow rate of component I through freeboard, kmol sec^{-1}
 F_g Flow rate of gas
 F_{gr} Flow rate of gas oil, kg sec^{-1}
 F_s Catalyst circulation rate, kg sec^{-1}
 F_{sr} Catalyst mass flow rate, kg sec^{-1}

H_{db} Height of dense bed, m
 H_{fb} Height of freeboard, m
 H_f Heat of reaction
 H_{oc} hydrogen content of coke
 H_{rab} Heat of reaction for cracking of lump a to lump b, KJ kg^{-1}
 H_i Bubble to emulsion heat interchange coefficient, kcal $m^{-2} sec K$
 K_{ab} Reaction rate constant for cracking of lump a to lump b, s^{-1}
 K_{bci} Bubble-to-cloud gas interchange coefficient for component I, sec^{-1}
 K_{mi} Bubble to emulsion gas interchange coefficient for component I, sec^{-1}
 K_g thermal conductivity of gas mixture, kcal m^{-1} , sec K
 M_A Molecular weight of air
 N Vortex component
 P Regenerator pressure, atm
 r_i Rate of reaction I in emulsion phase, kmol/ $m^3 sec$
 R_i Rate of reaction I in bubble phase, kmol/ $m^3 sec$
 T_e Emulsion phase temperature, $^{\circ}C$
 T_b Bubble phase temperature, $^{\circ}C$
 T_a Air inlet temperature, $^{\circ}C$
 T_f Freeboard temperature, $^{\circ}C$
 t_c Catalyst residence time, sec
 T_c Catalyst inlet temperature, $^{\circ}C$
 T_{ref} Reference temperature
 u Superficial velocity of emulsion gas with zero net flow of solids, $m sec^{-1}$
 U_i Inlet gas velocity
 U_c Superficial velocity of emulsion gas, $m sec^{-1}$
 U_{go} Gas superficial velocity, $m sec^{-1}$
 U_{br} Velocity of rise of a single bubble, $m sec^{-1}$
 U Superficial velocity of gas in the free board, $m sec^{-1}$
 U_{gb} Superficial velocity of bubble gas, $m sec^{-1}$
 U_{mf} Minimum Fluidisation Velocity, $m sec^{-1}$
 U_s Superficial velocity of solid feed, $m sec^{-1}$
 w_i Weight fraction of I
 x Dimensionless axial length
 Z height of the reactor, m

Greek Letters

ΔH_i Heat of Reaction I, kcal mol^{-1}
 ϵ_c Voidage of emulsion phase
 ϵ_f Voidage of free board
 δ Bubble fraction for dense bed
 ρ_g Density of the gas mixture, $kg m^{-3}$
 ρ_g Gas density
 ρ_s Density of the catalyst, $kg m^{-3}$
 ρ_w Water density
 η_i Efficiency of fraction I
 μ Gas viscosity

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