

EFFECT OF PRESSURE ON HEIGHT OF REGENERATOR DENSE BED IN AN FCCU

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Abstract

A fluid catalytic cracker unit converts heavy hydrocarbon petroleum streams into more valuable, lighter hydrocarbon fractions. During this process coke is deposited on the catalyst surface. This catalyst is regenerated in a regenerator which has two regions- the dense bed and dilute phase. The effect of regenerator pressure on dense bed height is studied by simulating an empirically tuned model in the present study.

Keywords: Fluid catalytic cracker; Regenerator; dense bed; regenerator pressure.

1. Introduction

Fluid catalytic cracking (FCC) unit plays most important role in the economy of a modern refinery that it is use for value addition to the refinery products. Because of the importance of FCC unit in refining, considerable effort has been done on the modelling of this unit for better understanding and improved productivity. in last sixty years, the mathematical modelling of FCC unit have matured in many ways but the real process whose hardware is ever-changing to meet the needs of petroleum refining is characterized by complex interactions among feed quality, catalyst properties, unit hardware parameters and process conditions.

In fluidized catalytic cracking processes, high molecular weight hydrocarbon liquids and vapours are contacted with hot, finely-divided, solid catalyst particles, either in a fluidized bed reactor or in an elongated transfer line reactor, and maintained at an elevated temperature in a fluidized or dispersed state for a period of time sufficient to effect the desired degree of cracking to lower molecular weight hydrocarbons of the kind typically present in motor gasoline and distillate fuels.

In the catalytic cracking of hydrocarbons, some non-volatile carbonaceous material or coke is deposited on the catalyst particles. Coke comprises highly condensed aromatic hydrocarbons. As coke accumulates on the cracking catalyst, the activity of the catalyst for cracking and selectivity of the catalyst for producing gasoline blending stocks diminishes. Catalyst which has become substantially deactivated through the deposit of coke is continuously withdrawn from the reaction zone. This deactivated catalyst is conveyed to the stripping zone where volatile deposits are removed with an inert gas or steam at elevated temperatures. The catalyst particles are then reactivated to essentially their original capabilities by substantial removal of the coke deposits in a suitable regeneration process. Regenerated catalyst is then continuously returned to the reaction zone to repeat the cycle. Fig.1 shows the schematic of FCC unit. In regenerator, catalyst is in fluidised state. We have two zones in fluidised bed, dense phase and dilute phase. Height of dense bed is an important factor while operating a regenerator.

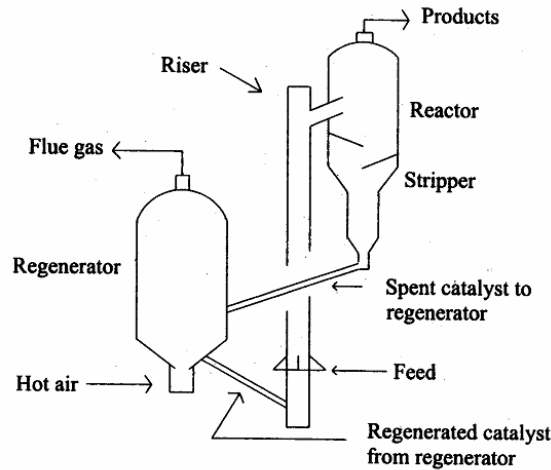


Figure 1. Schematic of FCC unit^[1]

Completeness of combustion of C to CO₂ depends not only on the temperature and air flow rate that exist in the dense bed, but also on the dense bed height. For example, too low a regenerator dense bed height can reduce the residence time and promote the occurrence of afterburning in dilute phase. The reason for this is that most of the catalyst is in dense bed, and therefore a significant heat sink has been removed. Thus, for the same amount of CO burning, the temperature rise is much greater, because the heat capacity of the flue gas with entrained catalyst is much lower than that of the flue gas which contains catalyst in dense bed. On the other hand, if dense bed height is high, it creates recirculation problems for regenerated catalyst.

Normally flow rate of air entering the regenerator is considered as the controlling parameter to decide the dense bed height. But the authors, by simulating the mathematical model of an Fluid Catalytic Cracker have observed that pressure also plays an important role in deciding dense bed height.

2. Validification of the FCCU model

2.1. Model Development

Riser Kinetics

Here I considered a five lumped kinetic scheme for optimization as shown in Fig.2.

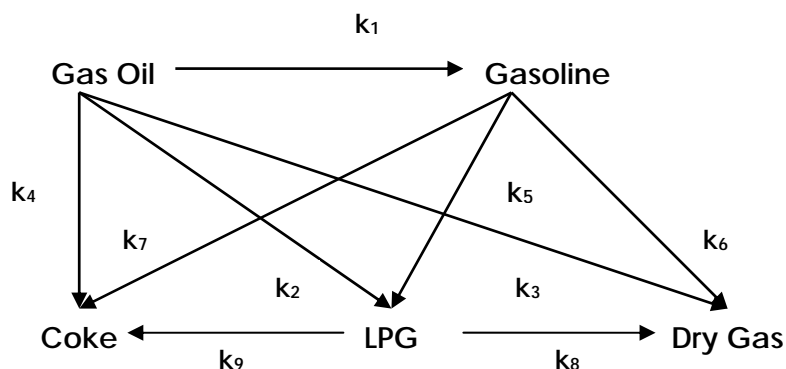


Fig. 2. Schematic diagram of the five-lump model examined.

Kinetic expression

For each reaction, a kinetic expression (r_i) was formulated as a function of product yield (y_i), deactivation function (ϕ) and kinetic constants (k_i). Gas oil cracking was considered as a second order reaction and gasoline and LPG as first order^[2]. The use of first order reaction for cracking of LPG has been discussed in the literature^[3]. The

exponential law was assumed for catalyst decay (ϕ). Based on these assumptions, the reaction rates of the model [4] are:

$$\text{Gas oil: } (r_1) = -(k_1 + k_2 + k_3 + k_4)y_1^2\phi \quad (1)$$

$$\text{Gasoline: } (r_2) = (k_1y_1^2 - k_5y_2 - k_6y_2 - k_7y_2)\phi \quad (2)$$

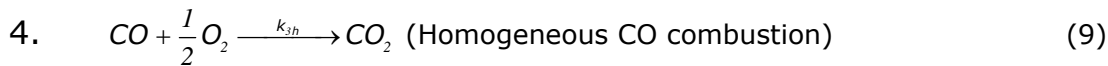
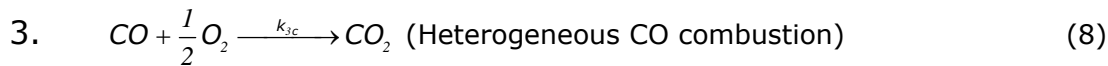
$$\text{LPG: } (r_3) = (k_2y_1^2 + k_5y_2 - k_8y_3 - k_9y_3)\phi \quad (3)$$

$$\text{Dry gas: } (r_4) = (k_3y_1^2 + k_6y_2 + k_8y_3)\phi \quad (4)$$

$$\text{Coke: } (r_5) = (k_4y_1^2 + k_7y_2 + k_9y_3)\phi \quad (5)$$

Regenerator Kinetics

The following combustion reactions are considered to be taking place in the regenerator.



The coke combustion in the reaction given by equation (6) and (7) are proportional to C_{rgc} and partial pressure of O_2 in the regenerator (P_{O_2}). The CO combustion reaction equations (8) and (9) are proportional to P_{O_2} and partial pressure of CO in the regenerator (P_{CO}). The rate expressions for these reactions are [5]:

Rate of reaction 1

$$r_1 = (1 - \varepsilon) \rho_{cat} k_1 \frac{C_{rgc} f_{O_2}}{M W_{coke} f_{Tot}} P_{rgn} \quad (11)$$

Rate of reaction 2

$$r_2 = (1 - \varepsilon) \rho_{cat} k_2 \frac{C_{rgc} f_{O_2}}{M W_{coke} f_{Tot}} P_{rgn} \quad (12)$$

Rate of reactions 3

$$r_3 = (x_{pt} (1 - \varepsilon) \rho_{cat} k_{3c} + \varepsilon k_{3h}) \frac{f_{O_2} f_{CO}}{f_{Tot}^2} P_{rgn}^2 \quad (13)$$

Where, x_{pt} is a relative catalytic CO combustion rate

$$\left(\frac{CO}{CO_2} \right)_{surface} = \frac{k_1}{k_2} = \beta_c = \beta_{c0} \exp\left(\frac{-E_\beta}{RT}\right) \quad (14)$$

If k_c is overall coke combustion rate then

$$k_c = k_1 + k_2 = k_{c0} \exp\left(\frac{-E_c}{RT}\right) \quad (15)$$

Where,

$$k_1 = \frac{\beta_c k_c}{\beta_c + 1} = \frac{\beta_c k_{c0} \exp\left(\frac{-E_c}{RT}\right)}{\beta_c + 1} \quad (16)$$

$$k_2 = \frac{k_c}{\beta_c + 1} = \frac{k_{c0} \exp\left(\frac{-E_c}{RT}\right)}{\beta_c + 1} \quad (17)$$

Riser Modelling

In the modelling of riser, the hydrodynamics characteristics of riser of a FCC plant are considered by using a plug flow model with radian dispersion.

In the model development of riser reactor, the commonly used assumptions are follows below:

- 1 At the riser inlet hydrocarbon feed comes into contact with the hot catalyst coming from the regenerator and instantly vaporizes (taking latent heat and sensible heat from the hot catalyst). The vapour thus formed move upwards in thermal equilibrium with the catalyst.
- 2 There is no loss of heat from the riser and the temperature of the reaction mixture (hydrocarbon vapours and catalyst) falls only because of the endothermicity of the cracking reactions.
- 3 Gas phase velocity variation on account of gas phase temperature and molar expansion due to cracking is considered. Ideal gas law is assumed to hold.
- 4 Heat and mass transfer resistances are assumed as negligible.
- 5 There are no radial temperature gradient in the gas and solid phase.
- 6 Conradson carbon residue of feed is zero.
- 7 Catalyst deactivation is non-selective and related to coke on catalyst only.
- 8 Gas oil cracking is a second order reaction but cracking of gasoline and LPG are first order reactions.
- 9 Dry gas produces no coke.
- 10 Heat capacities and densities are constant through the length of the reactor.

On the basis of above assumption, the mole balance for the j^{th} lump over a differential element can be written as follows^[4]

$$\frac{dF_j}{dh} = A_{ris} H_{ris} (1 - \varepsilon) \rho_c \sum_{i=1}^9 \alpha_{ij} r_i \quad j = 1, 2, \dots, 5 \quad (18)$$

Where $j = 1$ to 5 represents gas oil, gasoline, LPG, dry gas, and coke respectively. $i = 1$ to 9 represents the reactions as shown in fig. 3.1.

$$MW_g = \sum_{j=1}^5 x_j MW_j \quad (19)$$

$$\rho_v = \frac{P_{ris} MW_g}{RT} \quad (19)$$

$$\varepsilon = \frac{\frac{F_{feed}}{\rho_v}}{\frac{F_{feed}}{\rho_v} + \frac{F_{rgc}}{\rho_c}} \quad (20)$$

$$\alpha_{ij} = \frac{MW_i}{MW_j} \quad (21)$$

The molecular weights of different lumps used for the calculation of α_{ij} are given in Table. 3. The rate equation in (kmol)/(kg cat.) (s) are given by following expressions:

$$r_i = k_{0,i} \exp\left(-\frac{E}{RT}\right) C_i^2 \phi \quad \text{for } i = 1, 2, 3, 4 \quad (22)$$

$$r_i = k_{0,i} \exp\left(-\frac{E}{RT}\right) C_i \phi \quad \text{for } i = 5, 6, 7 \quad (23)$$

$$r_i = k_{0,i} \exp\left(-\frac{E}{RT}\right) C_i \phi \quad \text{for } i = 8, 9 \quad (24)$$

Where, C_1 , C_2 and C_3 are concentration of gas oil, gasoline and LPG respectively.

For the modelling of the catalyst deactivation the function ϕ was related to coke on catalyst as follows^[7]:

$$\phi = (1 + 51C_c)^{-2.78} \quad (25)$$

Due to the endothermic cracking reactions in the riser, there is a temperature drop along the height of the riser. The enthalpy balance across a differential element of height dh of the riser can be represented as follows^[4]:

$$\frac{dT}{dh} = \frac{A_{ris} H_{ris} (1 - \varepsilon) \rho_c \sum_{i=1}^9 r_i (-\Delta H_i)}{F_{rgc} C_{Pc} + F_{feed} C_{Pv}} \quad (26)$$

The regenerated catalyst and the preheated feed are mix at the base of the riser. Temperature at this zone can be determined from enthalpy balance. Assuming a 10°C drop in temperature of the regenerated catalyst during its journey in transfer line, the riser bottom temperature is calculated as follows^[4]:

$$T(h = 0) = \frac{F_{rgc} C_{Pc} (T_{rgn} - 10) + F_{feed} C_{Pfi} T_{feed} - \Delta H_{evp} F_{feed} - Q_{loss,ris}}{F_{rgc} C_{Pc} + F_{feed} C_{Pfv}} \quad (27)$$

where,

$$Q_{loss,ris} = 0.019 [F_{rgc} C_{Pc} (T_{rgn} - 10) + F_{feed} C_{Pfi} T_{feed} - \Delta H_{evp} F_{feed}]$$

To match the riser base temperature with the plant value, so the empirical term $Q_{loss,ris}$ can be incorporated.

Stripper Modelling

Due to the lack of required design and operating data the performance of the stripper has been idealized. The temperature drop across the industrial stripper was observed to be 8-12°C. So the temperature drop across the stripper was assumed to be 10°C^[8]. The temperature of the spent catalyst entering the regenerator is calculated by following equation:

$$T_{sc} = T_{ris,top} - \Delta T_{st} \quad (28)$$

Regenerator Modelling

The development of the steady state model is based on the following assumptions:

- 1 Gas is in the plug flow throughout the bed and in thermal equilibrium with surrounding bed.
- 2 Catalyst in dense bed is well mixed and isothermal with uniform carbon on the catalyst particles.
- 3 Kinetics of the coke combustion assumes catalyst particles to be 60 μm sizes.
- 4 Resistance to mass transfer from gas to catalyst phase is negligible.
- 5 Mean heat capacities of gases and catalyst are assumed to remain constant over the temperature range encountered.
- 6 All entrained catalyst is returned via cyclones.

Dense Bed Modelling

The spent catalyst from the reactor enters the regenerator dense bed in which coke is burn-off in the presence of air to CO, CO₂ and H₂O. The oxidation of hydrogen is assumed to be instantaneous and complete and hence the amount of oxygen available for the carbon burning reactions at the dense bed inlet is that remaining after the hydrogen combustion reaction.

Differential Balances:

Material and energy balance across a differential element of height dz of the dense bed are as follows^[4]:

Material Balance:

$$\frac{df_{O_2}}{dz} = -A_{rgn} \left(\frac{r_1}{2} + r_2 + \frac{r_3}{2} \right) \quad (29)$$

$$\frac{df_{CO}}{dz} = -A_{rgn} (r_3 - r_1) \quad (30)$$

$$\frac{df_{CO_2}}{dz} = -A_{rgn} (r_2 + r_3) \quad (31)$$

Energy Balance:

$$\frac{dT}{dz} = 0 \quad (32)$$

Initial Conditions:

$$f_{H_2O} = F_{rgc} (C_{sc} - C_{rgc}) \frac{C_H}{M W_H} \quad (33)$$

$$f_{O_2}(0) = 0.21 F_{air} - \frac{1}{2} f_{H_2O} \quad (34)$$

$$f_{CO}(0) = f_{CO_2}(0) = 0 \quad (35)$$

$$f_{N_2} = 0.79 F_{air} \quad (36)$$

Total gas flow rate at any cross section is given by:

$$f_{Tot} = f_{H_2O} + f_{O_2} + f_{CO} + f_{CO_2} + f_{N_2} \quad (37)$$

Bed Characteristics:

Gas molar density (kmol/m³)

$$\rho_g = \frac{P_{rgn}}{R T_{rgn}} \quad (38)$$

Superficial linear gas velocity (m/s)

$$u = \frac{F_{air}}{\rho_g A_{rgn}} \quad (39)$$

Void fraction was calculated using the correlations^[9]:

$$\varepsilon = \frac{0.305 u_1 + 1}{0.305 u_1 + 2} \quad (40)$$

Where, u_1 = superficial linear velocity in ft/s.

Dense bed height is also calculated using the correlation^[9]:

$$Z_{bed} = 0.3048(TDH) \quad (41)$$

$$TDH = TDH_{20} + 0.1(D - 20) \quad (42)$$

$$\text{Log}_{10}(TDH_{20}) = \text{log}_{10}(20.5) + 0.07(u_1 - 3) \quad (43)$$

Where, D is the regenerator diameter in ft, TDH means transport disengaging height.

The volume of a compartment j^{th} in the regenerator dense bed is given by:

$$V_j = A_{rgn} \Delta z_j \quad (44)$$

Where, $\Delta z_j = \frac{H_{densebed}}{N_c}$

Overall Balances:

Carbon balance for the regenerator in the dense bed^[4]:

$$C_{rgc} = \frac{F_{sc} C_{sc} (1 - C_H) - (f_{CO(Zbed)} + f_{CO_2(Zbed)}) M W_c}{F_{rgc} (1 - C_H)} \quad (45)$$

Heat Balance:

Applying heat balance across the regenerator dense bed gives the expression for the dense bed temperature^[4]:

$$T_{rgn} = T_{base} + \left\{ \frac{f_{CO(Zbed)} H_{CO} + f_{CO_2(Zbed)} H_{CO_2} + f_{H_2O} H_{H_2O} + F_{air} C_{P_{air}} (T_{air} - T_{base}) + F_{sc} C_{P_c} (T_{sc} - T_{base}) - Q_{loss,rgn}}{F_{rgc} C_{P_c} + f_{CO_2(Zbed)} C_{P_{CO_2}} + f_{CO(Zbed)} C_{P_{CO}} + f_{O_2} C_{P_{O_2}} + f_{H_2O} C_{P_{H_2O}} + f_{N_2} C_{P_{N_2}}} \right\} \quad (46)$$

Dilute Phase Modelling

Plug flow kinetics is assumed in the dilute phase. The main reaction taking place in the dilute phase is the oxidation of CO to CO₂. As a result both carbon concentration and temperature varies as a function of height in the dilute phase. Material and energy balance in the dilute phase results in the following equation^[4].

Material Balance:

$$\frac{df_{O_2}}{dz} = -A_{rgn} \left(\frac{r_1}{2} + r_2 + \frac{r_3}{2} \right) \quad (47)$$

$$\frac{df_{CO}}{dz} = -A_{rgn} (r_3 - r_1) \quad (48)$$

$$\frac{df_{CO_2}}{dz} = -A_{rgn} (r_2 + r_3) \quad (49)$$

$$\frac{df_c}{dz} = -A_{rgn} (r_1 + r_2) \quad (50)$$

Energy Balance:

$$\frac{dT_{dil}}{dz} = \frac{1}{f_{Tot} C_{P,Tot}} (H_{CO} \frac{df_{CO}}{dz} + H_{CO_2} \frac{df_{CO_2}}{dz}) \quad (51)$$

Where,

$$C_{P,Tot} = \left(\frac{C_{P_{N_2}} f_{N_2} + C_{P_{O_2}} f_{O_2} + C_{P_{CO}} f_{CO} + C_{P_{CO_2}} f_{CO_2} + C_{P_{H_2O}} + C_{P_c} F_{ent}}{f_{Tot}} \right)$$

Entrainment is calculated by using the correlation^[9]:

$$Y = \frac{W}{V \rho_f} \quad (52)$$

$$X = \frac{V^2}{g D_p \rho_p^2} \quad (53)$$

$$\log_{10} Y = \log_{10} 60 + 0.69 \log_{10} X - 0.445 (\log_{10} X)^2 \quad (54)$$

$$F'_{ent} = W A_{rgn} \quad (55)$$

$$F_{ent} = 0.4535 (F'_{ent}) \quad (56)$$

Catalyst density and void fraction in the dilute phase:

$$\rho_{dil} = \frac{F_{ent}}{A_{rgn} u_{rgn}} \quad (57)$$

$$\varepsilon = 1 - \frac{\rho_{dil}}{\rho_c} \quad (58)$$

The initial flow rate of coke in the dilute phase is given by:

$$f_c(0) = F_{ent} C_{rgc} \frac{(1 - C_H)}{12} \quad (59)$$

The height of dilute phase is calculated from the following expression:

$$Z_{dil} = Z_{rng} - Z_{bed} \quad (60)$$

2.2 Simulation of model

- 1 The initial values of coke on regenerated catalyst (**Crgc**) and regenerator temperature (**Trgn**) is assumed.
- 2 Rate equations of all the five lumps are integrated along the riser length with a small step size using **Runga Kutta** method and subsequently coke on spent catalyst (**Csc**) and riser top temperature (**Tris,top**) is calculated.
- 3 Then the dense bed calculations of regenerator are carried out to obtain calculated values of carbon on regenerated catalyst (**Crgc**) and dense bed temperature (**Trgn**).
- 4 Rate equations in the dilute phase bed are also integrated to obtain flue gas composition and temperature.
- 5 If the calculated values of (**Crgc**) and (**Trgn**) do not match with the assumed values then repeat the calculations with the calculated values as the new assumed values.
- 6 Gas oil conversion, product yields in the reactor and flue gas composition and temperature are evaluated at the converged values of (**Crgc**) and (**Trgn**).

Table. 1. Kinetic and Thermodynamic Parameters used for the Reactor Modeling [4]

Rate Constant	Reaction	Frequency Factor	Activation Energy (kJ/kmol)	Heat of Reaction (kJ/kmol)
k1	Gas Oil to Gasoline	19 584.55	57 540	45 000
k2	Gas Oil to LPG	3 246.45	52 500	159 315
k3	Gas Oil to Dry Gas	5 59.90	49 560	159 315
k4	Gas Oil to Coke	41.44	31 920	159 315
k5	Gasoline to LPG	65.40	73 500	42 420
k6	Gasoline to Dry Gas	0.00	45 360	42 420
k7	Gasoline to Coke	0.00	66 780	42 420
k8	LPG to Dry Gas	0.32	39 900	2 100
k9	LPG to Coke	0.19	31 500	2 100

Table. 2. Kinetic Parameters used for Regenerator Modelling [5]

Kinetic Parameter	Frequency Factor	Activation energy (E/R, K)
β_c	2512	6 795
K_c (1/(atm)(s))	1.069×10^8	18 890
K_{3c} ((kmol CO)/ (kg cat.)(s))	117	13 890
K_{3h} ((kmol CO)/ (m ³) (atm) ² (s))	5.07×10^{14}	35 555

Table. 3. Thermodynamics and other Parameters used for Simulation [4]

Parameter	Numerical Value
$C_{p,c}$ (kJ/kg K)	1.003
$C_{p,fl}$ (kJ/kg K)	3.430
$C_{p,fv}$ (kJ/kg K)	3.390
C_{p,N_2} (kJ/kg K)	30.530
C_{p,O_2} (kJ/kg K)	32.280
C_{p,H_2O} (kJ/kg K)	36.932
$C_{p,CO}$ (kJ/kg K)	30.850
C_{p,CO_2} (kJ/kg K)	47.400
ΔH_{evp} (kJ/kg)	350.0
H_{CO} (kJ/kg)	1.078×10^5
H_{CO_2} (kJ/kg)	3.933×10^5
H_{H_2O} (kJ/kg)	2.42×10^5
X_{Pt}	0.10
ρ_c (kg/m ³)	1089.0
C_H (kg H ₂ /kg coke)	0.165
D_p (ft)	2.0×10^{-4}
$MW_{gas\ oil}$	350
$MW_{gasoline}$	114
MW_{LPG}	52
$MW_{drv\ gas}$	30
MW_{coke}	12
Gas oil lump Boiling	>221
Gasoline lump	30-220

Table. 4. Design data for FCC unit [8]

Parameter	Value
Riser length (m)	37.0
Riser diameter (m)	0.68
Regenerator length (m)	19.34
Regenerator diameter (m)	4.52
Catalyst hold up in the regenerator (vol. %)	40.5

2.3 Comparison of simulated results with experimental data

Table.5. shows a comparison of model simulated yields with the measured values. The first seven rows of the table are required by the MATLAB code as inputs. The match between the model predicted yields and the measured data is good.

Table. 5. Comparison of the model predicted parameters with the plant value

Type	Set	
	Measured ^[4]	Simulated
Feed flow rate (kg/s)	31.47	-
Feed preheat temperature (K)	617.4	-
Catalyst flow rate (kg/s)	208.33	-
Riser pressure (atm)	2.457	-
Air flow rate (kmol/s)	0.56	-
Air preheat temperature (K)	490.3	-
Regenerator pressure (atm)	2.588	-
Riser top temperature (K)	765.5	764.389
Regenerator temperature (K)	930.2	930.78
Gas oil (%)	48.1	45.42
Gasoline (%)	32.6	34.71
LPG (%)	12.1	12.46
Dry gas (%)	3.1	3.37
Coke (%)	4.1	4.04
Dense bed height (m)	-	6.452
Coke on regenerated catalyst	-	0.55726×10^{-4}
Entrained catalyst flow rate	-	32.609

Since the simulated values of the model match well with the experimental values, the authors used this model for studying the effect of total pressure in regenerator on height of dense bed. To study the effect the authors varied the pressure of regenerator from 2.3atms to 2.8atms while keeping the air flow rate constant at 0.56 kmol/s.

3. Results and discussion

From Fig.3. it is observed that increasing the pressure of regenerator decreases the dense bed height of the regenerator. While simulating it was also observed that by increasing the pressure there was negligible increase in the conversion of CO to CO₂. The increase was negligible may be due to the fact that the increase in pressure which favours conversion of CO to CO₂ is counter affected by the decrease in dense bed height.

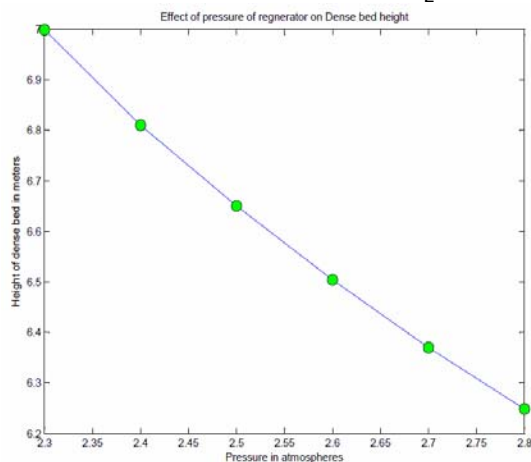


Figure 3. Effect of pressure of regenerator on dense bed height

But in practical conditions reduced bed height will decrease the proper mixing of air with catalyst, also the residence time of flue gas in dense bed decreases, as a result combustion may not be proper and CO concentration in flue gas leaving the dense bed may be high.

As the height of dense bed decreases, the length of dilute phase increases, as a result residence time of flue gas in dilute phase increases and chances of CO converting to CO₂, which may cause after-burn effects. On the other hand if we decrease the pressure the height of dense bed increases, which cause problems in circulating the regenerated catalyst. Therefore optimum maintenance of regenerator pressure is of great importance for proper functioning of a fluid catalytic cracker unit.

4. Conclusion

A mathematical model of fluid catalytic cracker was simulated and simulation results were compared with experimental data. Using this model effect of regenerator pressure on regenerator dense bed height was studied and it was found that along with air regenerator pressure also plays an important role in determining the dense bed height of the regenerator.

NOMENCLATURE

A	cross-sectional area of regenerator, ft^2
A_{rgn}	cross-sectional area of regenerator, m^2
A_{ris}	cross-sectional area of riser, m^2
C_c	coke on catalyst at any location, (kg of coke) (kg of catalyst) $^{-1}$
C_i	concentration of i th lump, kmol m^{-3}
C_H	weight fraction of H_2 in coke, (kg of H_2) (kg of coke) $^{-1}$
C_{rgc}	coke on regenerated catalyst, (kg of coke) (kg of cat) $^{-1}$
C_{sc}	coke on spent catalyst, (kg of coke) (kg of cat) $^{-1}$
C_{pc}	heat capacity of catalyst, $\text{kJ kg}^{-1} \text{K}^{-1}$
C_{pfl}	heat capacity of liquid feed, $\text{kJ kg}^{-1} \text{K}^{-1}$
C_{pfv}	heat capacity of vapor feed, $\text{kJ kg}^{-1} \text{K}^{-1}$
C_{pi}	mean heat capacity of i [H_2O , N_2 , O_2], $\text{kJ kg}^{-1} \text{K}^{-1}$
C_{ptot}	heat capacity of (total) mixture, $\text{kJ kg}^{-1} \text{K}^{-1}$
D	diameter of regenerator, ft
D_p	average diameter of catalyst particle, ft
E_c, E_β	activation energies, kJ kmol^{-1}
E_i	activation energy of i th reaction, kJ kmol^{-1}
f_i	molar flow rate of i [CO , CO_2 , H_2O , N_2 , O_2 , carbon] in the regenerator, kmol s^{-1}
f_{tot}	total gas flow rate at any location in the regenerator, kmol s^{-1}
F_{air}	flow rate of air feed to the regenerator, kmol s^{-1}
F_{ent}	entrained catalyst flow rate, kg s^{-1}
F_{feed}	feed flow rate of oil, kg s^{-1}
F_j	molar flow rate of j th lump, kmol s^{-1}
F_{rgc}	flow rate of regenerated catalyst, kg s^{-1}
F_{sc}	flow rate of spent catalyst, kg s^{-1}
g	gravitational acceleration, 32.2 ft s^{-2}
h	dimensionless height of riser ($\equiv z/H_{\text{ris}}$)
ΔH_{evp}	heat of vaporization of gas oil feed, kJ kg^{-1}
H_i	heat of formation of i , kJ kmol^{-1}
ΔH_i	heat of i th reaction, kJ kmol^{-1}
H_{ris}	height of riser, m
$k_{0,i}$	frequency factor for i th reaction
k_c	overall rate of combustion of coke
k_i	reaction rate constant for i th reaction
MW_c	molecular weight of coke, kg kmol^{-1}
MW_g	average molecular weight of gas phase, kg kmol^{-1}
MW_j	molecular weight of j th lump, $j = 1, 2, \dots, 5$, kg kmol^{-1}
MW_H	molecular weight of H_2 , kg kmol^{-1}
P_{rgn}	pressure in regenerator, atm
P_{ris}	pressure in riser, atm
r_i	rate of the i th reaction, $i = 1-9$ (riser); $i = 10-12$ (regenerator), $\text{kmol (kg catalyst)}^{-1} \text{s}^{-1}$ or $\text{kmol m}^{-3} \text{s}^{-1}$
R	universal gas constant, $\text{J K}^{-1} \text{kmol}^{-1}$
T_{air}	temperature of air fed to the regenerator, K
T_{base}	base temperature for heat balance calculations, K (assumed, 866.6 K)
T_{dil}	temperature of dilute phase at any location, K
T_{feed}	temperature of gas oil feed, K
T_{rgn}	temperature (uniform) of dense bed, K
T_{ris}	temperature of riser at any location, K

$T_{ris, top}$	temperature at top of riser, K
T_{sc}	temperature of spent catalyst [= $T_{ris, top} - \Delta T_{st}$], K
ΔT_{st}	temperature drop in stripper (assumed 10 K)
u	velocity of gas in the riser or the regenerator, $m s^{-1}$
u_1	superficial linear velocity, $ft s^{-1}$
W	catalyst entrainment flux, $lb (ft^2 \text{ regenerator area})^{-1} s^{-1}$
x_j	mole fraction of j th lump, $j = 1, 2, \dots, 5$
x_{pt}	relative (catalytic) CO combustion rate
Y	(kg catalyst entrained in dilute phase) (kg fluidizing vapor) $^{-1}$
z	height from the entrance of the regenerator, m
Z_{bed}	height of the dense bed, m
Z_{dil}	height of the dilute phase, m
Z_{rgn}	total height of the regenerator, m

Greek Letters

α_{ij}	stoichiometric coefficient of j th species in i th reaction, based on mass
β_c	CO/CO ₂ ratio at catalyst surface in regenerator
ε	void fraction in riser or regenerator at any location
ε_{dil}	void fraction in the dilute phase at any location
ρ_c	density of solid catalyst (not including void fraction), $kg m^{-3}$
ρ_{den}	density of catalyst in the dense bed, $kg m^{-3}$
ρ_{dil}	density of catalyst in the dilute phase, $kg m^{-3}$
ρ_f	density of fluidization vapor, $lb ft^{-3}$
ρ_g	density of gas phase in the regenerator, $kmol m^{-3}$
ρ_p	density of catalyst particle (solid), $lb ft^{-3}$
ρ_v	density of vapor at any location, $kg m^{-3}$
φ	activity of the catalyst

Subscripts

i, j i^{th} or j^{th} lump (1, gas oil; 2, gasoline; 3, LPG; 4, dry gas; 5, coke)

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