Fluid catalytic cracking process engineering essay

INTRODUCTION

Fluid catalytic cracking process, which is now more than 60 years old, is the cornerstone of most of the petroleum refineries. It has proven to be the most-efficient process available for the conversion of gas oils and residue into more valuable lighter hydrocarbons. Many refiners consider the catalytic cracking process to be the highest profit generating unit in the entire refinery. In earlier times, Fluid Catalytic Cracking Unit (FCCU) was operated broadly in two modes, they are;

Maximum gasoline mode

Maximum distillate mode

But with the advent of Reformulated gasoline (RFG), these are now operated in maximum olefin mode. FCCU is a very sophisticated unit with many factors affecting each other and the overall process. In some processes investigation of factors impact is done by changing one factor at a time while keeping other factors constant. In case of FCCU it is almost practically impossible to obtain a clear indication; as, change in one single factor leads to change(s) in one or more other factors. This whole phenomenon is a natural consequence of the " heat balance" of FCCU. If the unit is to operate at steady state, then the unit has to be in heat balance condition. At this stage the heat requirement in the reactor is satisfied by burning coke in the regenerator and transferring the energy to the reactor through circulating hot catalyst. Heat balance around the reactor-regenerator can be used to predict the effects of process changes although the exact degree of the changes may be difficult to establish. It is one step at a time thought process and rather difficult to pin down exact numbers without a careful study of yields and coke laydown rates as affected by changing variables. In this work a plant data is taken as reference and based on that, calculations have been done to find out the net heat of endothermic reactions occurring in the riser reactor, assuming that the unit is operating at steady state and that the riser is an isothermal one. Then as per the products slate, a 7-lumped model is considered from various literatures and based on the kinetics of reactions, rate equations are formed and with the knowledge of available kinetic parameters the differential temperature drops along the height of the riser are calculated.

PROCESS DESCRIPTION

More than a dozen types of FCCU are operating worldwide. But the basic designs of all these type remain the same. FCCU comprises of two parts;

Riser reactor, in which catalytic cracking reactions occur

Regenerator, in which burning of coke (deposited during cracking) from the catalytic sites is done

Figure 1 shows a schematic diagram of a typical FCCU. The feed is preheated in a furnace and

(Figure: 1- Schematic Diagram of a typical FCCU)

injected at the bottom of the riser along with a small amount of steam. This steam helps in dispersion of feed, good atomization and reduces coke formation by decreasing the partial pressure of hydrocarbon vapours. The feed is subsequently vaporized when it comes in contact with the hot catalyst from regenerator. The hydrocarbon vapours so formed undergo endothermic cracking reactions on their way up through the riser. The expansion of product vapours occurs through the length of the riser and the gas velocity increases with decreasing gas density. Hot catalyst particles provide the sensible heat and latent heat requirements for vaporizing the liquid feed and also endothermic heat of reaction for the cracking reactions. After a certain distance from the entry zone of the riser, the liquid feed is completely vaporized. Cracking reactions continue with the vapours moving up in the riser and the temperature is dropped along the length of the riser due to endothermic nature of cracking. The catalytic cracking is started and also completed in a very short period of time inside the riser reactor in which the catalyst is pushed upward by incorporating steam at various locations along the length of the riser and hydrocarbon vapours. Mixture of catalyst and hydrocarbon vapour travels up in the riser into the reactors. Steams injected at different locations in the riser are as follows,

Fluffing steam at the bottom of the riser

Dispersion steam along with fresh feed injectors

Riser dilution steam above the fresh feed injectors

Dispersion steam along with recycle stream injectors

Aeration steam into the riser “ J” bend to fluidize the catalyst

Along with this some other locations are there where steam is injected. They are as follows;

Spent catalyst standpipe aeration steam

Regenerated catalyst standpipe aeration steam

Reactor quench steam

Reactor dome steam

Post riser quench steam

Stripping steam into strippers

Mixture of catalyst and hydrocarbon vapour is discharged from the riser to the riser cyclone assembly. The bulk of the spent catalyst is separated from product vapours in the cyclone assembly. If necessary the vapours leaving the riser cyclones are routed into secondary cyclone assembly located inside the reactor vessel. Separated catalysts flow through each cyclone dip leg into the stripper. Product vapours leave the reactor cyclones and flow into the main fractionator through the reactor overhead vapour line. Quench steam is injected inside the reactor vessel to reduce the temperature, so as to minimize post riser thermal cracking reactions and coke formation. Reactor dome steam is provided to sweep hydrocarbons and avoid dead areas on top of the reactor vessel that may lead to thermal cracking and coking in that area. The separated catalyst from the riser and reactor cyclone assemblies enters the catalyst stripper.

As the catalyst flows down the stripper, it gets stripped off the entrained hydrocarbon vapours by the up flowing steam. Stripping enhances the product recovery and reduces the carryover of hydrocarbon to the regenerator along with the spent catalyst thereof. Fluffing steam ensures the fluidization of the circulating catalyst. Stripped catalyst from the stripper flows into the regenerator dense bed through the spent catalyst standpipe (SCSP). Catalyst level in the stripper is maintained by spent catalyst slide valve (SCSV). Aeration steam is provided in the SCSP to ensure proper flow and fluidization of spent catalyst.

Coke adsorbed on the spent catalyst during cracking reaction is been removed in the regenerator by burning off the coke with air. Air is supplied from the air blower to the regenerator through multiple distributors. Air is also introduced at different locations of the regenerator, they are as follows;

T-grid air

Regenerated catalyst standpipe (RCSP) hopper aeration air

RCSP aeration air

Regenerator fluffing air at the bottom near the “ J” bend

The regenerator can be operated in two modes;

Partial combustion mode

Complete combustion mode

For partial combustion mode, a CO boiler is needed to convert CO to CO_2. The current discussion is for complete combustion mode regenerator.

Flue gas from the regenerated dense bed flows to the two stage regenerator cyclone assembly. Here the entrained catalyst is separated from the flue gas. The separated catalyst flows back to the dense bed through cyclone dip legs. Flue gas from the cyclone flows out from top of the regenerator through a flue gas line. Total air flow to the regenerator is regulated based on the desired level of oxygen in flue gas. Too low O2 concentration will cause coke build up on regenerated catalyst and CO release from regenerator. Too high O2 concentration will lead to regenerator cooling. So, regenerator flue gas is regularly examined for O2, CO, CO2, NO2, SO2 analysis.

FEED CHARACTERIZATION

The only constant in FCC operation is the frequent change in feedstock quality. That’s why two feeds with similar boiling point ranges can exhibit huge differences in cracking performance and product yields. Feed characterization is one of the most important activities in monitoring the FCC process. Feed characterization is the process of determining physical and chemical properties of the feed. Understanding feed properties and also knowing their impact on unit’s performance is an essential thing. Trouble shooting, catalyst selection, unit optimization and subsequent process evaluation, all depend on feedstock. Feed characterization relates product yields and qualities to feed quality. Analytical techniques like mass spectrometry are sophisticated and not practical for determining complete composition of FCC feedstock. Simpler empirical correlations are often used. They are as follows;

oAPI gravity and UOP K

Boiling range

Average boiling point

Carbon residue

Metals

Sulphur, Nitrogen and Oxygen

oAPI gravity and UOP K

It is a specific gravity relating the density of oil to the density of water. The empirical formula for this is;

oAPI – 131. 5 (3. 1)

Feed to an FCC can range from 15o to 45o API. If the API gravity increases the charge stock will crack more readily and for the same reaction temperature there will be greater conversion. Secondly at a constant conversion level, there will be greater gasoline yield with slightly lower octane.

A rough indication of the quantities of paraffin present is a characterization factor which relates boiling point to specific gravity, is called the UOP K factor. This is given by;

(3. 2)

Where:

CABP = cubic average boiling point, oR

SG = specific gravity at 60 oF

Higher the UOP K value more is the paraffinic nature of the feedstock.

Boiling Range

The boiling range of FCC feed varies from an initial point of 500oF to an endpoint of about 1000oF. There are two boiling point ranges which are used to describe the lighter material in the feed. They are;

Per cent over 430oF

Per cent over 650oF

The first quantifies the amount of gasoline in the feed. The second one quantifies the light fuel oil in the charge.

Average boiling point

Average boiling point of the FCC feed depends on the average molecular weight. An increase in average boiling point and molecular weight will typically cause the following;

The charge will crack more readily, so at constant reactor temperature conversion will increase

At constant conversion, yield of C4 and lighter will decrease

Olefinic content of the product will decrease

Regenerator temperature will tend to rise

At constant conversion, the gasoline yield will increase about 1% for an increase in the molecular weight of 20.

Carbon residue

The carbon residue of a feedstock is an indirect measure of its coke producing nature. Values may be determined by either Conradson or Ramsbottom methods. The carbon residue can be a useful number for determining possible contamination in storage. Entrainment in vacuum tower is a common cause of increased carbon residue. Colour may be used to approximately evaluate the carbon content of the feedstock. Darker stocks tend to have higher carbon residues.

Metals

Organometallic compounds in the FCC feed can cause serious overcracking if the metals deposit on the catalyst. The cleanliness of a chargestock is given by a metals factor:

$Fm = Fe + V + 10 (Ni + Cu)$ (3. 3)

Where:

Fm = Metals Factor

Fe = Iron concentration

V = Vanadium concentration

Ni = Nickel concentration

Cu = Copper concentration

All metal concentrations are ppm by weight in the feed. A factor of 1. 0 is considered safe, over 3. 0 indicate a danger of poisoning of catalyst.

Sulfur, Nitrogen, Oxygen

Sulfur is as undesirable in FCC feed as it is in the feed to most of the refining units, causing corrosion of the equipment and increased difficulty in treating products. At 50% conversion about 35% sulfur charged is converted to H2S, and at 70% conversion the figure will rise to 50%. Nitrogen produces NH3 and CN- in the reactors, and NOx and trace quantities of NH3 in the regenerator. These NH3 and CN- cause plugging and corrosion, while the NOx and NH3 in the flue gas cause environmental problems. Gas oil will absorb oxygen in storage unless the tanks are gas blanketed. This oxygen will combine with the compounds in the oil at about 450oF to form gum, which fouls heat exchangers.

FCC REACTION CHEMISTRY

Cracking reactions are predominantly catalytic, but some non-selective thermal cracking reactions do take place. The two processes proceed via different chemistry. The occurrence of both the reactions is confirmed by distribution of products. Catalytic cracking proceeds mainly via carbenium ion intermediates. There are three dominant reactions in cracking are catalytic cracking, isomerization, hydrogen transfer. The idealized reaction classes are tabled below with specific reactions to support them.

(Table: 1 – idealized reactions of importance in FCCU)

Reaction classes

Specific reactions

Cracking

n-$C_{10}H_{22}$ n-C_7H_{16} + C_3H_6 ; 1-C_8H_{16} 2C_4H_8

Hydrogen transfer

4C_6H_{12} 3C_6H_{14} + C_6H_6 ; cyclo-C_6H_{12} + 3 1-C_5H_{10} 3n-C_5H_{12} + C_6H_6

Isomerization

1-C_4H_8 trans-2-C_4H_8 ; n-C_6H_{10} iso-C_4H_{10} ; o-$C_6H_4(CH_3)_2$ m- $C_6H_4(CH_3)_2$

Transalkylation

C_6H_6 + m- $C_6H_4(CH_3)_2$ 2$C_6H_5CH_3$

Cyclization

1-C_7H_{14} CH_3-cyclo-C_6H_{11}

Dealkylation

Iso-C_3H_7-C_6H_5 C_6H_6 + C_3H_6

Dehydrogenation

n-C_6H_{14} 1-C_6H_{12} + H_2

Polymerization

3C_2H_4 1-C_6H_{12}

Paraffin alkylation

1-C_4H_8 + iso-C_4H_{10} iso-C_8H_{18}

Some of the reactions are endothermic in nature and some are exothermic in nature. Each reaction has a heat of reaction associated with it. The overall heat of reaction is the combination of both the types of heat of reactions. Though there are a number of exothermic reactions, then also the net reaction is endothermic. It is apparent that the type and magnitude of reactions have an impact on the heat balance of the unit. If the catalyst is with less hydrogen transfer characteristics, it will cause the net heat of reaction to be more endothermic. This in turn results in higher catalyst circulation and possibly a higher coke yield to maintain the heat balance.

FCC UNIT MATERIAL BALANCE

For this, a complete set of commercial plant data is used. The data is given in subsequent tables below;

FEEDSTOCK

(Table: 2 – Properties of feed components)

Feed

Unit

Hydrotreated

VGO

Un-hydrotreated

VGO

Light Coker Naphtha

Quantity, TMTPA

3200

800

170

% of total feed

wt%

76. 74

19. 18

4. 08

Density @ 15oC

gm/cc

0. 894

0. 932

0. 6762

CCR

wt%

0. 1

1. 2

—

Sulfur

wt%

0. 1

3. 32

0. 434

Hydrogen content

wt%

13

–

–

Ni + V

wppm

1

6. 38

–

Nitrogen

wppm

500

1594

30

ASTM Distillation, vol.%

D-1160, oC

D-1160, oC

D-86, oC

IBP

366

349

36

5

374

379

—

10

385

394

43

30

420

435

49

50

443

468

57

70

485

508

65

90

545

556

75

95

576

573

—

FBP

620

609

86

Bromine no.

107. 86

Paraffins

vol.%

46. 7

Olefins

vol.%

43. 38

Naphthenes

vol.%

7. 25

Aromatics

vol.%

2. 68

RON, clear

79. 4

Diene value

5. 31

WATSON K

12. 436

MW

82. 001

PRODUCT YIELDS

(Table: 3- product yields, Ex-reactor and Perfect fractionator basis)

Products

wt %

Weight (lbs. /hr.)

H2S

0. 39

4309

Hydrogen

0. 041

606

Methane

1. 06

11710

Ethane

1. 54

17010

ethylene

1. 76

19442

Dry gas

4. 401

48768

Propane

2. 86

31592

Propylene

9. 66

106708

n-butane

1. 69

18668

i-butane

5. 52

60976

butenes

7. 47

82516

LPG

27. 2

300460

LCN

14. 50

160174

MCN

23. 40

257978

HCN

3. 90

43082

LCO

16. 45

181713

CLO

4. 75

153347

COKE

5. 01

——-

OPERATING CONDITIONS

(Table: 4- Operating conditions for the Unit)

Riser-Reactor	Unit	Value
Fresh heavy feed rate (VGO)	m3/hr.	533. 4
Fresh light feed rate (Coker naphtha)	m3/hr.	30. 2
CLO recycle	m3/hr.	46
Riser top temperature	oC	540
Riser top pressure	Kg/cm2	1. 5

Feed preheat temperature

oC

350

Regenerator

Air to regenerator (dry basis)

Nm3/hr.

310717

Regenerator pressure

Kg/cm2

1. 9

Dense bed temperature

oC

640

Dilute bed temperature

oC

654

Flue gas temperature

oC

657

Blower discharge temperature

oC

226

Stripper

Stripping steam rate

Kg/hr.

5000

Stripping steam temperature

oC

290

Stripping steam pressure

Kg/cm2

10. 5

Base temperature

oC

0

Ambient temperature

oC

35

Flue gas composition

MW= 30. 6

O2

vol. %

2. 49

CO

vol. %

0. 005

CO2

vol. %

15. 58

N2

vol. %

81. 83

SO2

vol. %

0. 085

SO3

vol. %

0. 01

Now using the above data, amount of oxygen that was consumed by burning the hydrogen in coke is estimated. All the gas calculations are based upon 100 moles of flue gas. The oxygen consumed for H2O is given by the expression;

O2 consumed = * (vol. % of N2 in flue gas) – 2 * (vol. % of O2 in flue gas)

– 2 * (vol. % of CO2 in flue gas) – (vol. % of CO in flue gas) (5. 1)

So, O2 consumed = * (81. 83) – 2 * (2. 49) – 2 * (15. 58) – (0. 005)

= 7. 36

The weight of the hydrogen and carbon in the coke are calculated;

Weight = 2. 016 * (7. 36) + 12. 01 * (15. 58+0. 005)

= 202. 01

The temperature differentials are calculated; (oF basis)

Î” TRR = (Regenerator dense bed temperature – Riser outlet temperature) (5. 2)

= 1184 – 1004

Î” TRR = 180

Î” TRB = (Regenerator fluegas temperature – Blower discharge temperature)

(5. 3)

= 1215 – 439

Î” TRB = 776

Î” TRS = (Riser outlet temperature – Stripping steam temperature) (5. 4)

= 1004 – 554

Î” TRS = 450

The weight combined feed ratio is calculated as;

(Flow rate)CLO * (Density)CLO * 2. 204

CFR = (5. 5)

(Flow rate)Fresh feed * (density)fresh feed * 2. 204

=

CFR = 0. 074

The stripping steam and inert gases carried to the reactor by the regenerated catalyst are calculated on a weight per pound fresh feed basis;

Steam = (5. 6)

Steam = 0. 01

Inert gases = (5. 7)

Inert gases = 0. 007

The amount of hydrogen in the coke is calculated as;

Hydrogen in Coke, wt % = [2. 016 * 7. 36 / 202. 01] * 100 %

= 7. 35 wt. %

The air to coke ratio is;

Air to coke, wt/wt = (2897/202. 01) * (81. 83/79)

Air to coke, wt/wt = 14. 85 lbs air / lb coke

Where;

2897 is the molecular weight of air multiplied by 100 (basis of 100 moles of flue gas)

The weight of coke per hour may be calculated as;

Weight of coke, lbs/hr. = (4591) * 193. 23 / 14. 85

= 59738. 6 lbs/hr.

Where;

(310717 Nm3/hr. = 5178. 62 Nm3/min. = 193. 23 MSCFM

4591 = air rate conversion factor from MSCFM to lbs/hr.)

So, weight % of coke is then;

wt. % coke = * 100%

= (59738. 6 / 1104941. 7) * 100 %

wt. % coke = 5. 41

In the product yield table, the coke wt. % is indicted as 5. 01 wt%. But it is calculated as 5. 41 wt. %. Now the overall weight balance is as follows;

OVERALL WEIGHT BALANCE

INPUT:-

= Fresh feed + Coker naphtha + CLO recycle

= {(533. 4 * 0. 8 * 894 * 2. 204) + (533. 4 * 0. 2 * 932 * 2. 204)} + (30. 2 * 676. 2 * 2. 204) + (46 * 808 * 2. 204)

= 1186860. 1 lbs. / hr.

OUTPUT:-

= Total product yields + coke

= 1149831 + 59738. 6

= 1209569. 6 lbs. / hr.

So, error in weight balance is calculated as;

= INPUT – OUTPUT

= (1186860. 1 – 1209596. 6) lbs. / hr.

= – 22736. 5 lbs. / hr.

= – 1. 88 wt. %

= 98. 12 % closure

Now combustion heat of coke is determined as follows; (at hottest temperature = flue gas temperature = 1215oF)

Î” Hcomb = [(X) (vol. % of CO in flue gas) + (Y) (vol. % of CO_2 in flue gas) + (Z) (vol. % of O_2 consumed)] / (weight if hydrogen and carbon in coke) (5. 8)

= [(48000) * (0. 005) + (169743) * (15. 58) + (106472) * (7. 36)] / 202. 01

Î” Hcomb = 16971. 8 Btu / lb coke

Where;

X = heat of combustion of CO at 1215oF

Y = heat of combustion of CO_2 at 1215oF

Z = heat of combustion of H_2O at 1215oF

There is correction factor for the hydrogen in coke, this is given as;

Correction factor, C = 1133 – (134. 6) (wt. % hydrogen) (5. 9)

= 1133 – (134. 6) (7. 35)

= 143. 7

The net heat of combustion after using the correction factor is;

-Î” HC = 16971. 8 + 143. 7 Btu / lb coke

-Î” HC = 17115. 5 Btu / lb coke

At this point the reactor and regenerator heat balances are calculated. The catalyst supplies the heat to the reactor. The regenerator heat balance is calculated first using a basis of one pound of coke at the hottest regenerator temperature. The reactor heat balance is based on one pound of fresh feed.

HEAT BALANCE

REGENERATOR HEAT

(Figure: 2- Regenerator heat In – Out scheme)

HEATREG = Î” HCOMB. – Î” HCOKE – Î” HAIR – Î” HRADIATION LOSS (6. 1)

Now, Î” HCOKE = heat required to raise coke to combustion temperature

= (0. 4) * (Î” TRR) (6. 2)

Î” HAIR = heat required to raise air to combustion temperature

= (lb air / lb coke) * (0. 26) * (Î” TRB) (6. 3)

Î” HRADIATION LOSS = 250 Btu / lb coke

So, HEATREG = 17115. 5 – {(0. 4) * (180)} – {(14. 85) * (0. 26) * (776)} – 250

HEATREGHEATREG = 13797. 4 Btu / lb coke

-Î” HCSo, regenerator efficiency = *100% (6. 4)

= 80. 6

REACTOR HEAT

(Figure: 3- Reactor heat In – Out scheme)

HEATRX = Î” HFRESH FEED + Î” HRECYCLE + Î” HSTRIPPING STEAM + Î” HREACTION + Î” HRADIATION LOSS + Î” HINERTS (6. 5)

Î” HFRESH FEED, Î” HRECYCLE = heat required to raise fresh feed & recycle to reactor temperature

Î” HSTRIPPING STEAM = heat required to raise steam to reactor temperature

= Î” TRS * (0. 485) * (lb steam / lb fresh feed) (6. 6)

Î” HRADIATION LOSS = 2 Btu / lb fresh feed

Î” HINERTS = heat of inert gases carried from regenerator to reactor by regenerated catalyst

= Î” TRR * (-0. 275) * (lb inerts / lb fresh feed) (6. 7)

HEATRX = (enthalpy of fresh feed at riser outlet temperature – enthalpy of fresh feed at preheat temperature) + CFR (enthalpy of recycle feed at riser outlet temperature – enthalpy of recycle feed) + Î” TRS * (0. 485) * (lb steam / lb fresh feed) + 2 Btu / lb fresh feed + Î” TRR * (-0. 275) * (lb inerts / lb fresh feed) + Î” HREACTION

= (745 – 460) + 0. 074 * (745 – 460) + 450 * (0. 485) * 0. 01 + 2 + 180 * (-0. 275) * 0. 007 + Î” HREACTION

HEATRX = 310 + Î” HREACTION

Note:-

Enthalpies for the fresh feed and the recycle feed were calculated by taking respective UOP K values, oAPIs and the temperatures from the API technical data book.

Regenerator heat is calculated on a one lb of coke basis. This can be converted to one lb of fresh feed by use of weight % of coke term.

So, HEATRX = HEATREG () (6. 8)

Î” HREACTION + HEATRX = HEATREG () + Î” HREACTION (6. 9)

Î” HREACTION = HEATREG () + Î” HREACTION + HEATRX (6. 10)

But HEATRX = + Î” HREACTION

Putting this relation in equation (6. 10), the equation changes to

Î” HREACTION = HEATREG () –

Î” HREACTION = 13797. 4 * – 310

Î” HREACTION = 436. 44 Btu / lb fresh feed

So, HEATRX = 310 + 436. 44

HEATRX = 746. 44 Btu / lb fresh feed

(0. 275) (Î” TRR)Cat / Oil (wt. / wt.) = HEATRX (6. 11)

Cat / Oil (wt. / wt.) = 15 lb Catalyst / lb Oil

Catalyst circulation rate = (Cat / Oil) * (lb fresh feed / hr.) (6. 12)

= 15 * 1104941. 8

CCR = 16574127 lbs. / hr.

= 7524 MT/ hr.

Overall heat flow scheme for the whole FCCU can be shown as below;

(Figure: 4- Typical FCCU heat balance scheme)

Now, the net total endothermic heat of reaction is calculated through empirical formulae. But we took the assumption as the riser is an isothermal one. Practically it is not isothermal. The temperature at the base of the riser is higher than what is at the top of the riser or at the riser outlet. This is because the cracking reactions occurring along the length of the riser is endothermic in nature. So heat is being absorbed during the reaction and causes the temperature at that particular location to decrease. Gradually the temperature decreases and at the riser outlet the temperature is dropped significantly. In this context we can estimate the riser base temperature using empirical relations and therefore can estimate the drop in temperature at the next differential element up in the riser [DNS]. But before this a multi-lumped model is to be considered along with possible reaction schemes and there kinetic parameters.

SEVEN LUMP KINETIC MODEL

For this purpose a seven lump kinetic model proposed by Mehran Heydari et al. (2010) is used. They divided the model into seven lumps namely; VGO/Coker Naphtha, Clarified Oil, Light Cycle Oil, gasoline (LCN, MCN, and HCN), LPG, Dry gas and Coke. The schematic flow diagram is as follows;

(Figure: 5- Seven lump kinetic model in FCCU)

In order to develop a mathematical model for this particular system, certain assumptions has to be taken, they are as follows;

The riser is an one dimensional ideal plug flow reactor with no radial and axial dispersion

Reactor is an adiabatic riser

Feed viscosity and heat capacities of all components are constant

Fluid flow is not affected by the coke deposition on the catalyst

Feed is vaporized instantaneously in the riser entrance

All cracking reactions are taking place in the riser

The model considers seven lumps and eighteen reactions and eighteen kinetic constants. Molecular weights of different lumps and boiling ranges are given [DNS] in the table below;

(Table: 5- molecular weights and boiling ranges of lumps)

j

Lump

Molecular weight

(Kg/ Kmol)

Boiling range

(oC)

1

VGO

418. 7

349 – 620

2

CLO

291

232 -567

3

LCO

226

170 – 392

4

GASOLINE

114

30 – 228

5

LPG

65

—

6

DRY GAS

30

—

7

COKE

12

—

Values of kinetic constants and activation energies along with heat of reactions for each reaction are given in the table below ([DNS], [Mehran Heydari], [Praveen ch. & shishir sinha]);

(Table: 6- reaction schemes with kinetic parameters)

Reactions

Rate constants

(m3/ kg cat. hr.)

Activation energy

(KJ/Kmol)

Heat of reaction

(KJ/Kg)

VGO CLO

14. 93

50. 73

45. 821

VGO LCO

5. 78

50. 73

79. 213

VGO GASOLINE

11. 69

50. 73

92. 335

VGO LPG

3. 59

16. 15

159. 315

VGO DRYGAS

0. 35

16. 15

159. 315

VGO COKE

11. 55

16. 15

159. 315

CLO LCO

5. 78

50. 73

56. 314

CLO GASOLINE

0. 94

46. 24

128. 571

CLO LPG

0. 135

59. 75

455. 185

CLO DRYGAS

0. 0135

59. 75

455. 185

CLO COKE

0. 3272

59. 75

455. 185

LCO GASOLINE

0. 5742

46. 24

93. 030

LCO LPG

0. 0086

59. 75

704. 93

LCO DRYGAS

0. 0009

59. 75

704. 93

LCO COKE

0. 0596

59. 75

704. 93

GASOLINE LPG

0. 0003

78. 49

372. 10

GASO DRYGAS

0. 0001

78. 49

372. 10

LPG DRYGAS

0. 0033

59. 75

32. 30

The riser model is assumed to be a two phase model