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SHORT COMMUNICATION

Simulation and Optimization of Industrial FCC Units Using a 4-Lump Kinetic Model

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Abstract: Most of the previous work on the simulation and the bifurcation behavior of industrial FCC units involved a 3-lump kinetic model with coke and light hydrocarbon gases contributing one of the lumps with the ratio between them computed empirically from industrial data. In this study a 4-lump kinetic model is incorporated. This permits the calculation of the rate of coke formation thus decreasing the degree of empiricism in the model. The data from six different industrial units is used in the simulation. It was found that the model results are in good agreement with the corresponding plant data. The model predicts that the units are operated at the statically stable high temperature steady state whereas the use of the 3-lump kinetic model in previous work predicted that the units operate at the middle unstable steady state. This shows the importance of the structure of the kinetic model used for the determination of the stability characteristics of these important industrial units. The model is used to find the optimum operating conditions and the optimum performance is compared with the industrial performance.

Introduction

Fluid catalytic cracking (FCC) units are very important units in the petroleum refining industry. The major function of the unit is the conversion of heavy (large molecular weight) hydrocarbons into light (small molecular weight) hydrocarbons such as gasoline. Its basic mode of operation involves the cracking of the heavy constituents in the reactor using a circulating catalyst. The catalyst which loses activity in the reactor due to carbon deposition is reactivated in the regenerator by burning-off the carbon using air. The cracking reactions are endothermic whereas the regenerator reactions are exothermic. Therefore, the catalyst, as it circulates, serves the dual purpose of catalyzing and supplying the necessary heat to the reaction.

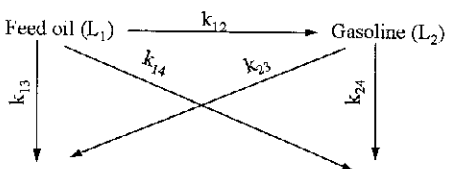
Various configurations of these units have been employed in the industry. In type IV FCC units part of the reaction takes place in the connecting tube (tube taking the regenerated catalyst from the regenerator to the reactor) and is completed in the reactor. However, in the riser configuration the majority of the reaction takes place in the connecting tube (the riser).

A large number of models are presented in the literature for the steady state simulation of fluid catalytic cracking (FCC) units [1-7]. Many of these models are based on a 3-lump kinetic model [3-5]. One of the most serious drawbacks of the 3-lump kinetic model is that it lumps the coke and the light hydrocarbon gases together even though they play completely different roles in the process. The amount of coke formed during the cracking is essential for the calculation of the amount of heat released during coke burning in the regenerator which in turn provides the heat required for the endothermic cracking reactions in the reactor. In the previous models the amount of coke was calculated from the ratio of coke to hydrocarbons gases obtained from plant data [4]. The introduction of a 4-lump kinetic model [8] separating coke from hydrocarbon gases, permits the calculation of deposited coke from the input data thus decreasing the degree of empiricism. Six sets of industrial data (from six different units from Canada, USA and Saudi Arabia) are used for checking the model. The data were divided into three groups (A = Canada units, B = USA units and C = KSA units) within each group there are two sets of data having the same catalyst, same feed, same reactor type, the two sets within each group differing only in operating conditions (see Table 1).

The Model

The Kinetic Model

The 4-lump kinetic model used is that of Lee *et al.* [8]. The four lumps are: feed oil, gasoline, hydrocarbon gases and coke as show below.



$$k_{ij} = k_{ijo}\, e^{-\frac{E_{ij}}{R.T}}$$

Gas oil cracking reactions are second order whereas gasoline cracking reactions are first order. The activation energies and frequency factors given by Lee et al. [8] are given in Table 2.

Table 1. Industrial data for the six industrial units

	U1 (Canada)	U2 (Canada)	U3 (USA)	U4 (USA)	U5 (KSA)	U6 (KSA)
Reactor type	riser reactor	riser reactor	riser reactor	riser reactor	Type IV	Type IV
Catalyst	zeolite	zeolite	zeolite	zeolite	Alumina/silica	Alumina/silica
Main product	naphta	naphta	light or heavy gasoline	light or heavy gasoline	Gasoline	Gasoline
Mol. wt. of fresh feed or product	310,100	310,100	400,105	400,105	180,110	180,110
Fresh feed flow rate, kg/s	72.566	72.945	87.594	86.247	16.782	13.476
Recycle flow rate, kg/s	15.581	12.206	5.881	4.887	2.109	2.111
Catalyst circulation rate, kg/s	606.311	633.346	985.0	801.988	88.605	117.1
Air flow rate, kg/s	47.803	45.884	69.119	69.609	10.67	10.67
Reactor pressure, KPa	241.1	243.93	204.8	212.7	225.49	225.49
Regenerator pressure, KPa	234.7	233.6	151.9	151.9	254.87	254.87
Fresh feed temperature, K	642.4	655	550	550	527.0	538.0
Air temperature, K	405	419	400.7	406.3	436.0	433.0
Mixed feed vaporization temperature, K	698	698	700	700	539.0	539.0
Heat of reaction, H_{12}, KJ/kg	180	180	100.4	100.4	125.5	125.5
Heat of reaction, H_{13}, KJ/kg	598.3	598.3	870.3	870.3	543.9	543.9
Heat of reaction, H_{14}, KJ/kg	891.2	891.2	1238.5	1238.5	627.6	627.6
Heat of reaction, H_{23}, KJ/kg	418.4	418.4	769.9	769.9	418.4	418.4
Heat of reaction, H_{24}, KJ/kg	711.3	711.3	1138.1	1138.1	627.6	627.6
Heat of vaporization of mixed feed, KJ/kg	255.2	255.2	185.78	185.78	264.86	264.86
Heat of coke formation, KJ/kg	33116.25	33116.25	36176.6	36176.6	45461.3	45461.3

Table 2. Base values of the pre-exponential factors and activation energies for the various reactions [8]

i,j	k_{ijo}	E_{ij} cal/mol.K
1,2	31064.07 m^3/kg.s	68281.9
1,3	177125.0 m^3/kg.s	89259.3
1,4	1465.98 m^3/kg.s	64606.1
2,3	0.904 1/s	52743.7
2,4	0.0221 1/s	115055.4
k_{ro}	$1.682x10^{11}$ kg/kg.s	28444.0

The heats of cracking depend on the type of feed and product and were calculated from the corresponding heats of formation of the different feed stocks and products (see Table 1). The coke burning rate and the heat of coke combustion have been obtained as in previous work [3,4].

Model characteristics

The model used here takes into consideration the two phase nature of the fluidized beds in both the reactor and regenerator. In the case of riser reactors, the flow pattern is certainly more complex than the two-phase formulation and the fluidization regime has been revised as suggested by Grace [9,10] to be applied to riser reactors. In this investigation and in order to elucidate the bifurcation behavior of the units, a pseudo empirical representation based on the industrially verified model, will be used for the riser reactor.

The model includes the dependence of all physical parameters on the state variables such as vapor densities and diffusivities which are functions of temperature and pressure. The change in the volumetric flow rates at inlet and outlet of both the reactor and regenerator has been accounted for. The model is totally dependent on input conditions except for the following output data:

1- **The H/C ratio in coke and the $CO_2/(CO+CO_2)$ ratio:** The heat of combustion of coke in the regenerator is calculated from an empirical correlation and is a function of H/C ratio in coke and the $CO_2/(CO+CO_2)$ ratio. These two ratios can be obtained from the flue gas composition available in plant data.

2- **The percentage of product recycled:** The amount of product depends on the operating temperature and is predicted by the model. The recycle stream is then calculated using the percentage recycled obtfrom plant data.

Model equation

The model assumptions and its detailed derivation are given elsewhere [3,4]. The definition of all variables and parameters are given in the notation section. The 4-lump kinetic scheme has been incorporated into the model and the following sets of equations have been derived:

Reactor

Feed oil mass balances:

Dense phase:

$$B(X_{1f} - X_1) = \varphi_R (X_1)^2 (\alpha_{12} e^{-\gamma_{12}/\gamma_R} + \alpha_{14} e^{-\gamma_{14}/\gamma_R}) \tag{1a}$$

Bubble phase:

$$X_{1b} = X_1 + (X_{1f} - X_1) e^{-\alpha_R H_R} \tag{1b}$$

Gasoline (product) mass balances:

Dense phase:

$$B(X_{2f} - X_2) = \varphi_R [(X_1)^2 (\alpha_{12} . e^{-\gamma_{12}/\gamma_R}) - X_2 . \alpha_{23} . e^{-\gamma_{23}/\gamma_R} - X_2 . \alpha_{24} . e^{-\gamma_{24}/\gamma_R}] \tag{2a}$$

Bubble phase:

$$X_{2b} = X_2 + (X_{2f} - X_2) e^{-\alpha_R H_R} \tag{2b}$$

Coke mass balance:

$$X_3 = \varphi_R . [(X_1)^2 . \alpha_{13} . e^{-\gamma_{13}/\gamma_R} + X_2 . \alpha_{23} . e^{-\gamma_{23}/\gamma_R}] \tag{3}$$

Light hydrocarbon gases mass balance:

Dense phase:

$$X_4 = \varphi_R . [X_1^2 . \alpha_{14} . e^{-\gamma_{14}/\gamma_R} + X_2 . \alpha_{24} . e^{-\gamma_{24}/\gamma_R}] \tag{4a}$$

Bubble phase:

$$X_{4b} = X_4 + (X_{4f} - X_4) e^{\alpha_R H_R} \tag{4b}$$

Heat balance:

Dense phase:

$$a_1(Y_G - Y_R) + B(Y_v - Y_R) + a_2(Y_f - Y_v) - (H_v + H_{LR}) + HRCB = 0 \quad (5a)$$

Bubble phase:

$$Y_{Rb} = Y_R + (Y_v - Y_R)e^{-\alpha_R H_R} \quad (5b)$$

Regenerator:

Mass balance on coke:

$$C_G(\varphi_G - \varphi_R) - C_B = 0 \quad (6)$$

Heat balance:

Dense phase:

$$B'(Y_{fa} - Y_G) + a_3(Y_R - Y_G) - H_{LG} + \beta_G . C_B = 0 \quad (7a)$$

Bubble phase:

$$Y_{Gb} = Y_G + (Y_{fa} - Y_G)e^{-\alpha_G H_G} \quad (7b)$$

The gasoline yield (Z_1) is defined as

$$Z_1 = \frac{(X_2 . G_d + X_{2b} . G_b)}{G_f} \quad (8)$$

The feed oil conversion (Z_2) is defined as:

$$Z_2 = \frac{(X_1 . G_d + X_{1b} . G_b)}{G_f} \quad (9)$$

Results and Discussion

In this section the results of the simulation and optimization of the six industrial units is presented. Because of some similarities and for space conservation in this paper, the graphical representation of the results are given for the first unit of group A, however the same analysis has been applied to all other units.

Simulation of the industrial units

The kinetic parameters of the 4-lump model [8] were derived for a specific catalyst and feed stock. The use of different catalysts and feed stocks necessitates some adjustment in the values of these parameters. Since each group of industrial data used

in this preliminary investigation pertains to the same catalyst and feed stock, adjustable parameters were introduced in the model to fit the first set of data in each group. The model was then used without any adjustable parameters and the results obtained were compared with the second set of plant data for the same group.

Tables 3-5 give a comparison between the plant data and the model results for the six units.

Table 3. Comparison between model results and plant data for the two industrial units of group A (Canada). ([a]) with adjustable parameters for the first industrial unit ([b]) without adjustable parameters for the second industrial unit

	Plant data unit # 1	Model results[a]	% error	Plant data unit # 2	Model results[b]	% error
dimensionless reactor temperature, Y_R	1.5982	1.5916	- 0.412%	1.6104	1.631	+1.279%
Dimensionless regenerator temperature, Y_G	1.956	1.9559	- 0.005%	1.9460	1.9951	+2.520%
Conversion/pass Z2	0.511	0.5075	- 0.680%	0.446	0.4591	+2.940%
Yield/pass Z1	0.3089	0.3069	- 0.647%	0.327	0.328	+0.310%
Coke, kg/s	4.4265	4.4486	+ 0.490%	4.946	4.637	-6.250%
LPG, kg/s	13.417	15.1397	+12.800%	14.414	15.581	+8.100%
Recycle, kg/s	15.581	15.730	+ 0.956%	12.206	12.637	+3.530%

Group A (Canada) units

Table 3 shows the comparison results for group A (Canada). Columns 2-4 are for first set in this group with adjustable parameters. The results of the comparison are excellent (within 1% deviation) for reactor dense phase temperature, regenerator temperature, conversion per pass, yield per pass, coke and recycle oil. The only exception is the LPG which showed an over estimation of 12.8%. The model is then compared to the second set of plant data of group A without any further adjustment of the parameters and the results are presented in the last 3 columns of Table 3. The predictions of the model (without adjustable parameters) are generally excellent with reactor and regenerator temperatures slightly over estimated (about 1.28% and 2.52% respectively). The model predicts slightly higher conversion per pass (about 2.94%), however the yield predicted by the model is quite close to that of the plant data (only 0.31% higher). The amount of coke formed is 6.25% less than the plant data while the LPG is 8.1% higher (better than LPG predicted by the model with adjustable parameters which was 12.8%). The recycle oil is slightly over predicted (3.53%).

Group B (USA) units

The comparison results for the two sets of group B (USA) are presented in Table 4. The percentage error between model predictions (with adjustable parameters) and plant data for the first set in this group are listed in columns 2-4 of Table 4. It is clear from these

columns that the model excellently predicted the reactor temperature (+0.063%), conversion per pass (0.0%) and recycle oil (0.051%). A fair agreement between model predictions and plant data is obtained for the yield per pass (+4.6%). The regenerator temperature, LPG and amount of coke are over estimated by 6.74%, 8.13% and 10.9% respectively. The last 3 columns of the table presents the results for the comparison of model predictions (without further adjustment of the parameters) and the plant data for the second set in group B. Excellent agreement has been obtained for LPG (+0.49%) and the dense phase reactor temperature (-1.26%). A fairly good agreement between model predictions and plant data for the second set has been obtained for yield per pass (-4.8%), regenerator temperature (+6.3%) and the conversion per pass (-7.8%). Both coke and recycle are over estimated by the model by 12.4% and 26.7% respectively.

Table 4. Comparison between model results and plant data for the two industrial units of group B (USA). ([a]) with adjustable parameters for the first industrial unit, ([b]) without adjustable parameters for the second industrial unit

	Plant data unit # 3	Model results[a]	% error	Plant data unit # 4	Model results[b]	% error
Dimensionless reactor temperature, Y_R	1.594	1.595	+ 0.063%	1.590	1.570	- 1.26%
Dimensionless regenerator temperature, Y_G	1.869	1.995	+ 6.740%	1.900	2.020	+ 6.30%
conversion/pass Z2	0.700	0.700	0.000%	0.770	0.710	-7.80%
Yield/pass Z1	0.400	0.4185	+ 4.600%	0.437	0.416	- 4.80%
Coke, kg/s	6.368	7.062	+10.900%	6.270	7.050	+12.40%
LPG, kg/s	22.250	24.060	+ 8.130%	24.020	24.140	+ 0.49%
Recycle, kg/s	5.882	5.879	- 0.051%	4.830	6.200	+ 26.7%

Group C (KSA) units

Table 5. gives the comparison results between model predictions and plant data for the two sets of group C (Saudi Arabia). The model with adjustable parameters (columns 2-4 of Table 5) excellently predicts the reactor temperature (0.0%), coke (0.12%), conversion per pass (-0.16%), yield per pass (1.37%), recycle oil (2.26%) and LPG (2.9%) for the first set of plant data in this group. Compared to the plant data, the model prediction is 9.9% higher for the regenerator temperature. The results of model predictions (without adjustable parameters) with the plant data for the second set of data in this group are shown in the last 3 columns of Table 5. The agreement between model predictions and plant data is excellent for reactor temperature (0.0%), regenerator temperature (1.8%) and yield per pass (2.16%). A good agreement is also obtained for the conversion per pass (-5.9%). The amount of LPG and coke are under predicted by the model by -11.7% and -14.4% respectively, while the recycle oil is over predicted.

Table 5. Comparison between model results and plant data for the two industrial units of group C (Saudi Arabia). (a) with adjustable parameters for the first industrial unit, (b) without adjustable parameters for the second industrial unit

	Plant data unit # 5	Model results [a]	% error	Plant data unit # 6	Model results [b]	% error
Dimensionless reactor temperature, Y_R	1.55	1.55	0.0%	1.57	1.57	0.0%
Dimensionless regenerator temperature, Y_G	1.91	2.1	+9.9%	1.886	1.92	+1.8%
Conversion/pass Z2	0.61	0.609	-0.16%	0.677	0.637	-5.9%
Yield/pass, Z1	0.4153	0.421	+1.37%	0.416	0.425	+2.16%
Coke, kg/s	0.8152	0.816	+0.12%	0.8425	0.721	-14.4%
LPG, kg/s	2.866	2.949	+2.9%	3.231	2.852	-11.7%
Recycle, kg/s	2.1089	2.1566	+2.26%	2.111	2.418	+14.5%

Figure 1 shows the heat function as a function of reactor output temperature and the corresponding yield of product and unreacted feed for the first unit of group A. It is clear from Fig. 1 and similar other figures (not shown) that the 6 units are operated in the multiplicity region and more specifically at the high temperature steady state.

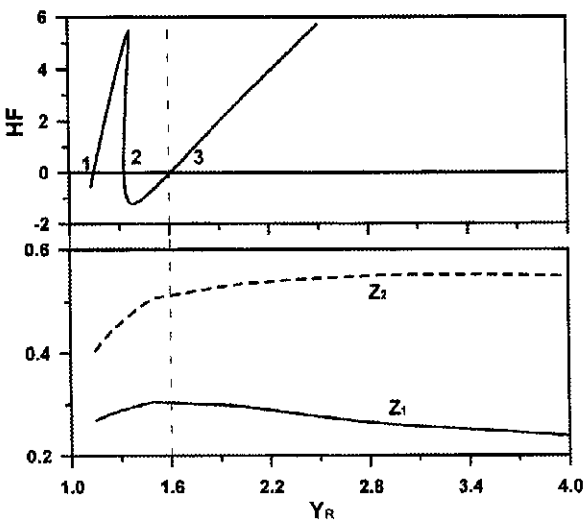


Fig. 1. Heat function curve (intersection with HF=0 line indicates steady states). Dashed vertical line show high temperature steady and the corresponding product yield (Z1) and conversion (Z2) for first unit of group A.

The low temperature steady state is always occurring at a reactor temperature below the vaporization temperature of the feed, i.e., no vaporization occurs and is therefore of no physical relevance; whereas the middle steady state occurs just above the vaporization temperature and is unstable.

Optimization

The effect of feed oil and air temperatures and flow rates on the performance of the six units has been investigated for all units in order to obtain a preliminary insight into the possible improvement of gasoline yield for these reactors through the manipulation of the operating variables. The effect of feed oil and air temperature is studied in the temperature range of [100 to 1000 K] which corresponds to a dimensionless range of [0.2,2] for the six industrial units. The range of investigation for the feed oil and the air flow rates is varied by [20% to 200%] of the respective industrial operating values for the six units. The results of the optimization exercise are summarized in Table 6, and the multiplicity of steady states regions are given in Table 7.

Table 6: Summary of the optimization results for the improvement of the % product yield

Unit	Y_{fo}	Y_f	FFR	AFR	% impro.
Group (A) unit # 1					
Industrial plant data	0.800	1.285	72.566	47.803	0.000
Simulation results	0.600	1.285	72.566	47.803	0.065
Simulation results	0.200	1.285	72.566	47.803	0.163
Simulation results	0.800	1.285	72.566	95.606	0.200
Group (A) unit # 2					
Industrial plant data	0.838	1.310	72.945	45.884	0.000
Simulation results	0.600	1.310	72.945	45.884	0.100
Simulation results	0.200	1.310	72.945	45.884	0.210
Simulation results	0.838	1.310	72.945	91.768	0.300
Group (B) unit # 3					
Industrial plant data	0.800	1.100	87.594	69.119	0.000
Simulation results	0.600	1.100	87.594	69.119	0.400
Simulation results	0.800	0.700	87.594	69.119	2.700
Simulation results	0.800	1.100	166.40	69.119	11.40
Simulation results	0.800	1.100	87.594	138.22	2.000
Group (B) unit # 4					
Industrial plant data	0.813	1.100	86.247	69.609	0.000
Simulation results	0.600	1.100	86.247	69.609	0.410
Simulation results	0.813	0.800	86.247	69.609	2.000
Simulation results	0.813	1.100	172.50	69.609	13.00
Simulation results	0.813	1.100	86.247	139.20	2.100
Group (C) unit # 5					
Industrial plant data	0.872	1.054	16.782	10.670	0.000
Simulation results	2.000	1.054	16.782	10.670	0.400
Simulation results	0.872	2.000	16.782	10.670	2.100
Simulation results	0.872	1.054	16.782	2.134	3.700
Group (C) unit # 6					
Plant data results	0.866	1.076	13.476	10.670	0.000
Simulation results	2.000	1.076	13.476	10.670	3.700
Simulation results	0.866	1.076	13.476	2.134	3.100

Table 7. Regions of multiplicity of steady states for the 6 industrial units

Unit	Air temp. K	Feed temp. K	FFR kg/s	AFR kg/s
#1 group A	100-1000	400-1000	14.513-145.13	9.561-95.61
#2 group A	100-1000	335-1000	14.588-145.88	9.177-91.77
#3 group B	100-1000	325-1000	17.519-175.19	13.82-138.2
#4 group B	100-1000	330-1000	17.249-172.49	13.92-139.2
#5 group C	100 - 760	100-1000	3.356-33.56	2.134-21.34
#6 group C	100 - 665	100-1000	2.695-26.95	2.134-21.34

U1, first unit in group A (Canada)

Figures 2-5 show the effect of air feed temperature, feed oil temperature, feed oil flow rate, and air flow rate respectively for the first unit in group A (Canada). Line 3 in these figures corresponds to the high temperature steady states (the plant is operating at this steady state, therefore the discussion is focused on this branch).

Effect of air feed temperature (Y_{fa})

Figure 2 shows the effect of air feed temperature (Y_{fa}) on the reactor performance for the first unit in group A (Canada). It is clear from the figure that both the reactor and regenerator temperatures increase with increasing air feed temperature. The product yield (Z_1) is improved by decreasing the air feed temperature. As listed in Table 6, an increase of 0.065% in the yield is achieved by decreasing the air feed temperature to Y_{fa}=0.6 instead of Y_{fa}=0.8 (plant condition). Further improvement (0.163%) in product yield is attainable at Y_{fa}=0.2. The conversion of feed oil (Z_2) increases with air feed temperature. The multiplicity region covers the wide range of dimensionless air feed temperature used [0.2-2.0] as clear from Fig. 2 (100 - 1000 K as listed in Table 7).

Effect of gas oil feed temperature (Y_f)

Figure 3 shows the effect of gas oil feed temperature (Y_f) on the performance of the first unit in group A. It is clear from the figure that below a feed temperature of Y_f= 0.8 quenching occurs with a drop in reactor temperature below that required for feed vaporization. The reactor temperature, regenerator temperature and feed oil conversion (Z2) increase significantly with feed oil temperature for $0.8 \leq Y \leq 1.4$ and they slightly increase for $Y_f > 1.4$. The product yield (Z1) increases only slightly with the decrease in feed temperature. The multiplicity region covers $0.8 \leq Y \leq 2.0$ (Fig. 3) i.e. $400 \leq T \leq 1000$ K as given in Table 7. At feed oil temperatures below Y_f=0.8 a quenched steady state exists as a unique solution as clear from Fig. 3.

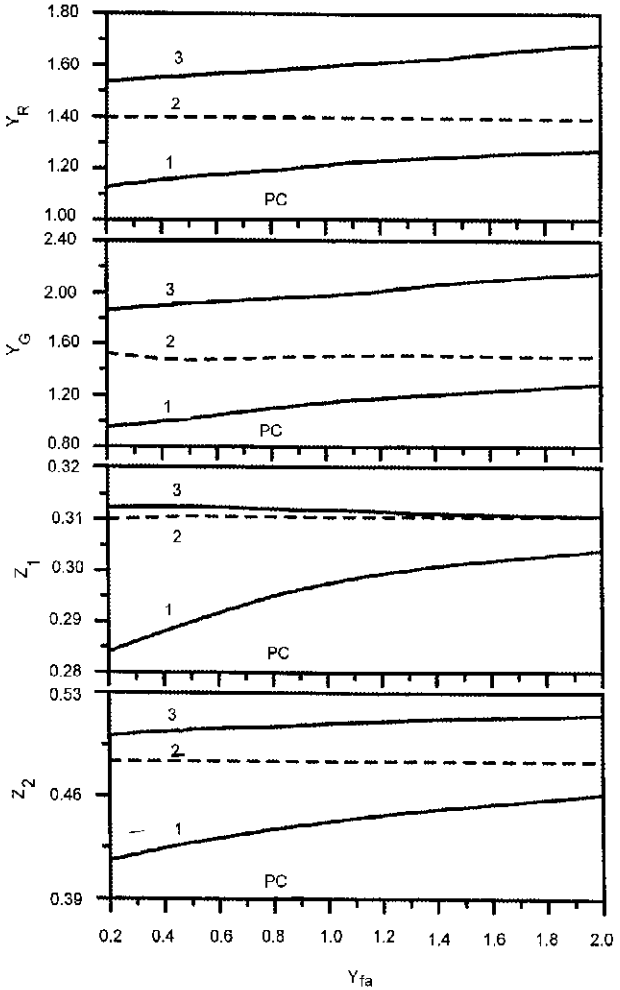


Fig. 2 Effect of air feed temperature (Y_{fa}) on reactor (Y_R) and regenerator (Y_G) temperature, product yield (Z_1) and conversion (Z_2) for first unit of group A. (1=stable low, 2=unstable middle and 3=stable high temperature steady states; PC= plant operating condition).

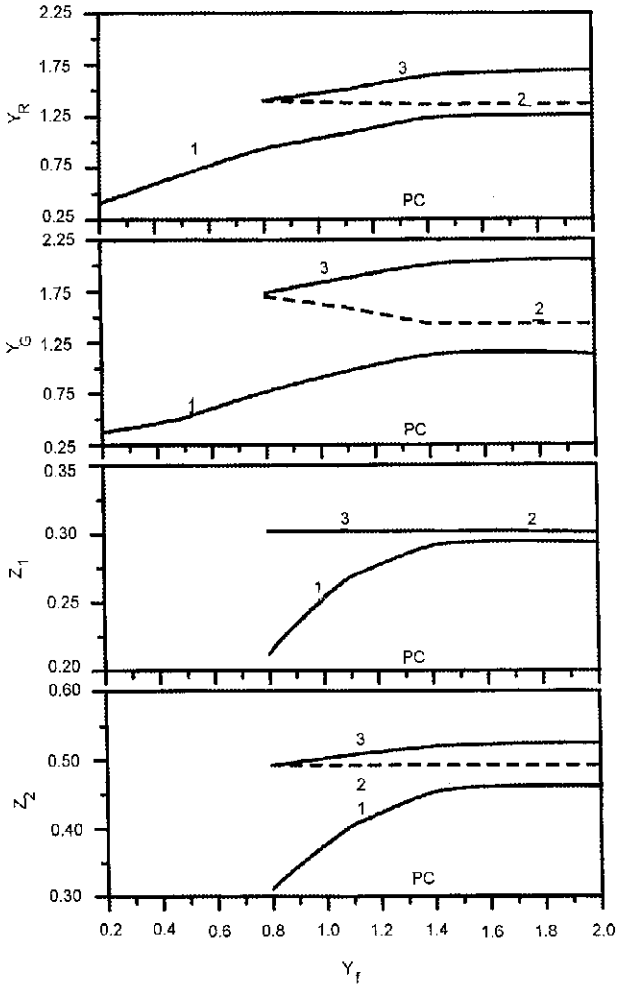


Fig 3. Effect of oil feed temperature (Y_f) on reactor (Y_R) and regenerator (Y_G) temperature, product yield (Z_1) and conversion (Z_2) for first unit of group A. (1=stable low, 2=unstable middle and 3=stable high temperature steady states; PC = plant operating condition).

Effect of feed flow rate

The effect of feed flow rate on the reactor performance is shown in Fig. 4. The reactor and regenerator temperature increase with increasing the feed flow rate, however, the conversion show an opposite behavior. The yield exhibits a maximum which corresponds with the operating flow rate, 72.566 kg/s (i.e. FFRF=1.0). The feed flow rate factor, FFRF, is the ratio between value of feed flow rate used in simulation and plant operating feed flow rate. The multiplicity region covers the wide range of investigation [14.513 to 145.13 kg/s] as given in Table 7.

Effect of air flow rate

Figure 5 presents the effect of air flow rate and it is clear from the figure that an increase in the air flow rate leads to a decrease in reactor, regenerator temperatures as well as conversion. By doubling the air flow rate from 47.803 to 95.606 kg/s (i.e. AFRF=2, the air flow rate factor, AFRF, is the ratio between value of air flow rate used in simulation and plant operating air flow rate) the yield increases by about 0.2% as shown in Table 6. The multiplicity region covers the wide range of investigation [9.56 to 95.6 kg/s] as shown in Table 7.

Similar results are obtained for the performance of the other 5 units (one unit in group A and two units in each of group B and C). The most important results for yield improvement are given in Table 6.

U2, second unit in group A (Canada)

For the second unit of group A, the product yield is improved by 0.1% and 0.21% when the air feed temperature is decreased from plant operating value (Y_{fa}=0.838) to Y_{fa}=0.6 and Y_{fa}=0.2 respectively. An increase in yield of 0.3% can be achieved by doubling the air flow rate of the industrial unit from 45.884 to 91.768 kg/s. As shown in Table 7, the multiplicity region covers the wide range of investigation for the air feed temperature and flow rate and for the feed oil flow rate. The multiplicity range for the feed oil temperature is [335 to 1000 K]with the low temperature branch (quenched steady state) existing as the only steady states branch for T_f<335 K.

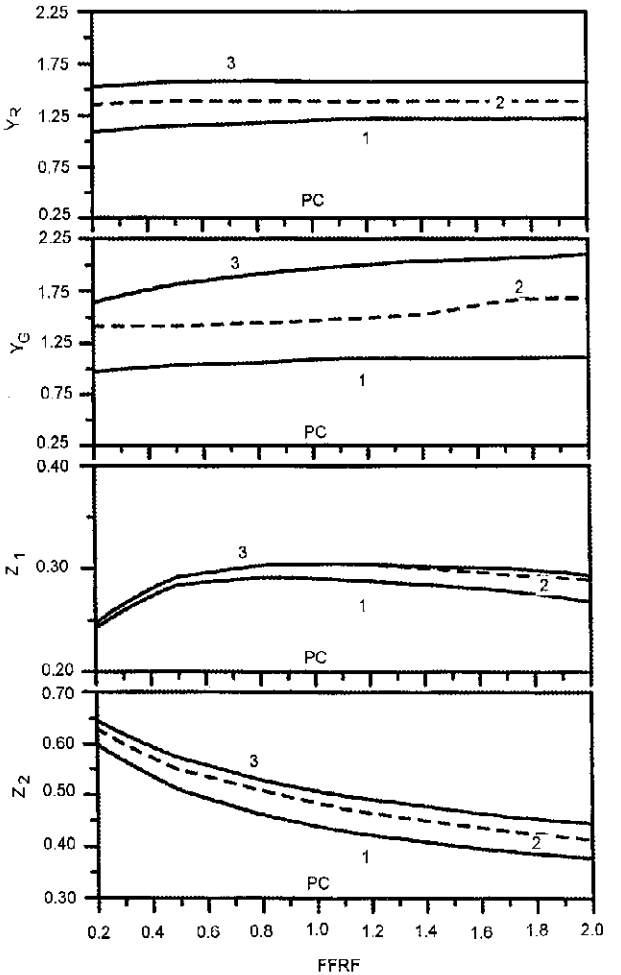


Fig. 4. Effect of feed flow rate (FFR) on reactor (Y_R) and regenerator (Y_C) temperature, product yield (Z_1) and conversion (Z_2) for first unit of group A. (1=stable low, 2=unstable middle and 3=stable high temperature steady states; PC= plant operating condition).

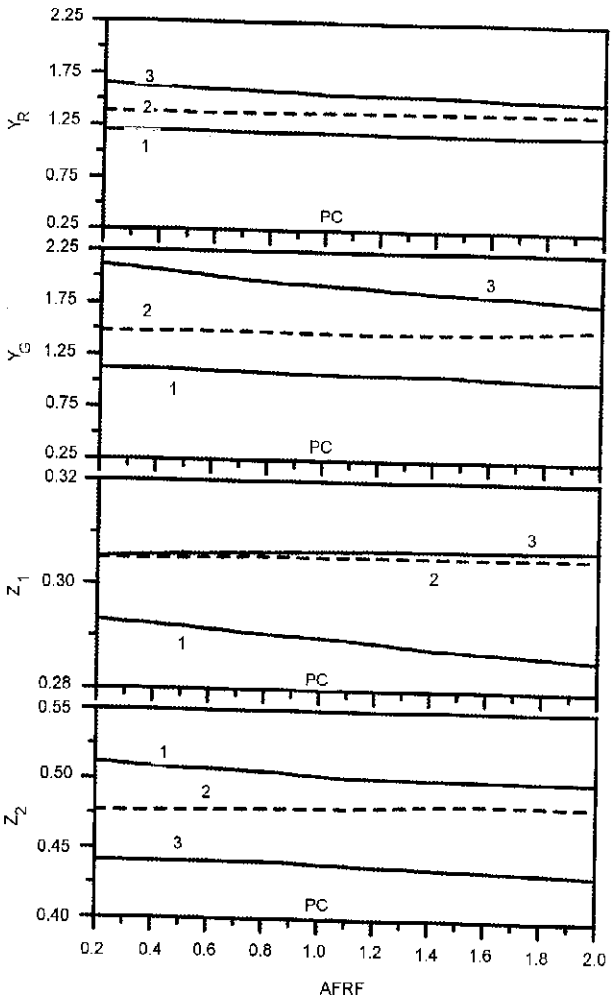


Fig. 5. Effect of air flow rate (AFR) on reactor (Y_R) and regenerator (Y_G) temperature, product yield (Z1) conversion (Z2) for first unit of group A. (1=stable low, 2 = unstable middle and 3=stable high temperature steady states; PC = plant operating condition).

U3, first unit in group B (USA)

For the first unit in group B (USA), the yield can be increased 0.4% by decreasing the air feed temperature from Y_{fa}=0.8 to Y_{fa}=0.6 (Table 6). A 2.7% improvement in the yield can also achieved by decreasing the feed oil temperature from Y_f=1.1 to Y_f=0.7. An increase in feed oil flow rate from 87.594 kg/s to 166.4 kg/s causes 11.4% increase in yield and by doubling the air flow rate of the industrial unit (from 69.119 to 138.238 kg/s) the yield increases by 2% (Table 6). The multiplicity region for this reactor cover the range of parameter values for air feed temperature, feed flow rate and air flow rate as shown on Table 7. However, quenching occurs for feed temperatures below 325 K (Y_f=0.65).

U4, second unit in group B (USA)

For the second unit in this group, quenching occurs when the feed temperature is below 330 K as shown in Table 7. The yield can be increased by 0.41% by decreasing the air temperature from 400 to 300 K and by decreasing the feed oil temperature from 550 to 400 K the yield increases by 2%. (Table 6). Doubling the feed oil flow rate from 86.247 to 172.494 kg/s) causes a yield increase of 13% whereas doubling the air flow rate of the industrial unit from 69.6 to 139.2 kg/s results in a yield increase of 2.1% (Table 6).

U5, first unit in group C (KSA)

For the first unit of group C (Table 6), a slight increase in the yield (about 0.4%) can be achieved by raising the air temperature from Y_{fa}=0.872 to Y_{fa}=2.0. However, when raising the oil feed temperature from Y_f=1.054 to Y_f=2.0, the yield increases by 2.1%. An increase of 3.7% in yield can be achieved by decreasing the air flow rate from 10.67 kg/s to 2.134 kg/s. Reactor ignition occurs for air feed temperature higher than 760 K (a unique high temperature steady state prevails) as given in Table 7.

U6, second unit in group C (KSA)

With regard to the second unit of group C (Table 6), a yield increase of 3.7% can be achieved by increasing the air feed temperature from Y_{fa}=0.866 to Y_{fa}=2.0. The yield can also be increased 3.1% achieved by decreasing the air flow rate from 10.67 kg/s to 2.134 kg/s.

Conclusion

A relatively simple mathematical model based on the two-phase theory of fluidization and using 4-lump kinetics is developed for the simulation and optimization of six industrial FCC units. The results predicted by the model simulates the

performance of the six industrial units quite accurately. The model predicts that the units are operated at their high steady state temperature. For the units of group A and B, the optimization exercise reveals that a slight improvement in product yield can be obtained by decreasing the air feed temperature or increasing the air flow rate, however significant improvement in the yield (11-14%) is achievable for the two units of group B if the feed flow rate is doubled. The yield in group C units can be improved by either increasing the air feed temperature (0.4-3.7%) or decreasing the air flow rate (3.1-3.7%).

The simple exercise presented in this work shows that considerable insight into the behavior of these industrially important units and optimization of their operating conditions for higher product yield can be achieved using a relatively simple model.

Notations

AFR	air flow rate, kg/s.
AFRF	air flow rate factor, dimensionless
a_1,a_2,a_3	dimensionless groups
B, B'	modified space velocities for the reactor and regenerator respectively, 1/s.
C_B	coke burning rate, kg coke burned/(kg solid.s)
FFR	feed flow rate, kg/s.
FFRF	feed flow rate factor, dimensionless.
G_f,G_d,G_b	volumetric flow rate in feed, dense and bubble phase respectively, m^3/s.
HF	heat function.
H_R,H_G	height of reactor and regenerator, respectively, m.
HL_R,HL_G	heat losses in reactor and regenerator respectively.
HRCB	dimensionless rate of heat absorption due to endothermic cracking reactions in the reactor.
Hv	dimensionless heat of vaporization for gas oil, kJ/kg.
k_{co}	pre-exponential factor for coke burning rate.
LPG	liquefied petroleum gas.
X_1	dimensionless dense phase oil concentration.
X_{1b}	dimensionless bubble phase oil concentration.
X_{1f}	dimensionless oil concentration in feed.
X2	dimensionless dense phase gasoline concentration.
X_{2b}	dimensionless bubble phase gasoline concentration.
X_{2b}	dimensionless gasoline concentration in feed.
X_3	dimensionless coke concentration.
X_4	dimensionless light gases concentration.
Y_f	dimensionless feed temperature.
Y_{fa}	dimensionless air feed temperature.

Y_G	dimensionless regenerator dense phase temperature.
Y_{Gb}	dimensionless regenerator bubble phase temperature.
Y_R	dimensionless reactor dense phase temperature.
Y_{Rb}	dimensionless reactor bubble phase temperature.
Y_V	dimensionless gas oil vaporization temperature.
Z_1	yield.
Z_2	conversion.

Greek letters

α_{ij}	dimensionless pre-exponential factors for the cracking reactions.
α_R, α_G	parameters in space velocity expression for reactor and regenerator respectively.
β_C	exothermicity factor for coke combustion reaction.
γ_{ij}	dimensionless activation energies for cracking reactions.
φ_R	catalyst activity within reactor.
φ_G	catalyst activity within regenerator.

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محاكاة و أمثلة وحدة تكسير صناعية ذات مهد مميع باستخدام معادلة كينتيكية رباعية

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ملخص البحث. استخدمت معظم الأبحاث السابقة لتمثيل ودراسة الظواهر التشعبية لوحدات التكسير الصناعية ذات المهد المميع معادلات كينتيكية ثلاثية وذلك بجمع الكربون و الغازات الخفيفة معا والحصول على نسبتهما من المعلومات الصناعية. تم ، في هذه الدراسة ، استخدام معادلة كينتيكية رباعية مما يسمح بحساب معدلات تكوّن الكربون. تمّ استخدام معلومات صناعية لست وحدات في المحاكاة و التمثيل ووجد أن نتائج النموذج جيدة مقارنةً بالمعلومات الصناعية. أظهر هذا النموذج أن الوحدات تعمل على المنحنى المستقر لدرجة الحرارة العالية في حين أظهر استخدام المعادلة الكينتيكية الثلاثية سابقًا أن الوحدات تعمل على المنحنى غير المستقر لدرجة الحرارة الوسطى. تظهر هذه النتيجة أهمية المعادلات الكينتيكية المستخدمة لتحديد مناطق الاستقرار في هذه الوحدات الصناعية المهمة. وتمّ أيضا استخدام النموذج لتحديد مناطق التشغيل الأمثل و مقارنة إنتاجية هذه المناطق بالإنتاج الفعلي للوحدات.