Modelling and Simulation of an Industrial Fluid Catalytic Cracking Unit

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To meet ever-changing market dynamics, industrial processes must adapt to new structures of the market demand and environmental restrictions. Using the model provided by Aspen HYSYS and adjusting the kinetic parameters, we aimed to simulate the functioning of an industrial plant as such to obtain results as close as possible to the values observed during operation. The adjusted model allows optimization of the industrial plant operation by extending the range of feeds or catalysts, and also by changing the plant geometry leading to higher yields of desired reaction products, which allows maximization of plant profitability.

Keywords: industrial FCC unit, kinetic model, Aspen HYSYS, simulation

Fluid catalytic cracking (FCC) is a versatile and often profitable unit of any modern refinery, which is used to convert heavy distillates and residual raw materials in lighter and more valuable products. The FCC unit has the greatest contribution to the production of the gasoline pool, but is also a significant provider of raw materials for the petrochemical industry, such as light olefins. Thus, the FCC unit has a major contribution to the overall profitability of the refinery [1].

To meet the ever changing requirements of the market, the plant operators often have to make adjustments in the operation parameters of industrial plants in order to improve yields and increase efficiency of the catalytic cracking unit. Major modifications, however, must be taken into account only after a concerted and successful effort that involves understanding the reactions chemistry, feed characteristics and performance of the equipment. In this respect, using a rigorous simulation models is of crucial importance [2].

Coupling hydrodynamic and kinetic models allow calculating, more or less detailed, the performance of an industrial plant. Modelling is therefore important to assess the main limiting phenomena and offer the opportunity to optimize reactor performance by adjusting reactor design and operating conditions so as to limit the steps to maximize the performance of the reactor. Independent on the level of detail involved in the modeling, it is essential to describe all relevant phenomena that must be accounted for, which is not necessarily an easy task [3].

Regarding fluid catalytic cracking modeling, the literature reports are divided into two categories, namely kinetic, and unit-level models, respectively. Kinetic models focus on chemical reactions taking place inside the riser or reactor section of the FCC unit and attempt to quantify the feed as a mixture of chemical entities to describe the rate of reaction from one chemical entity to another. In contrast, unit-level models contain several sub-models to take into account the integrated nature of a modern catalytic cracking unit. A basic unit-level model contains sub-models for riser / reactor, regenerator and catalyst transfer sections [2].

There are many models in the literature that reproduce the kinetics of reactions occurring in the riser/reactor system of the catalytic cracking unit. Simple models, with only few components, are suitable for specific simulations where the kinetic model is developed for a specific feed and a certain catalyst [4, 7]. A 6-lumps model was proposed in 1998 [8], and a similar one few years later (2002) [9]. These simple models describing the kinetics of cracking reactions of 3, 4, 5 and 6 specific chemical lumps have the disadvantage that some of them can not forecast separately different specific catalytic cracking products.

More complex models using a larger number of components can also be found in the literature. The most popular one is the model with 10 components proposed in 1976 [10]. An upgraded version of this model was proposed in 1987 [11]. This last model has 12 components, two gaseous components in addition to the original Jacobi model.

There are even more complicated models, which have advantages over previous models containing fewer components, namely that only one group of kinetic constants can be used for various feeds, and most of the reaction products can be estimated separately. As disadvantages, these models have a larger number of kinetic parameters to be estimated and an increased complexity of the possible numerical solutions, because each component needs one more differential equation in the mathematical model [12].

The most recent models published in the literature were 8 lumps [13] and 9 lumps models [14], and a model with 12 lumps [15]. However, besides the models published in the literature, there are models that lay at the foundation of commercial software packages that simulate industrial processes, as the Aspen Engineering Suite. Aspen's HYSYS existing kinetic model for catalytic cracking is based on a series of sub-models that can simulate the operation of the entire cracking unit. The thermal balance is satisfied both for riser and regenerator. The model is complete and includes the feed supply, the riser – reactor system, stripper, regenerator, valves for feed vaporization and cyclones [16]. The simplified scheme is suitable for obtaining quantitative estimates of fractionated reaction products from the reactor effluent. The model also includes several coke formation pathways and a distribution scheme for the feed sulfur and nitrogen into the reaction products.

The riser conversion kinetics is derived from the Mobil ten-lump mechanism [16]. The Aspen FCC model has expanded the number of reactant/product lumps to 21, and changed the functionality of several key lumps. All reactions are based on the well-understood first order
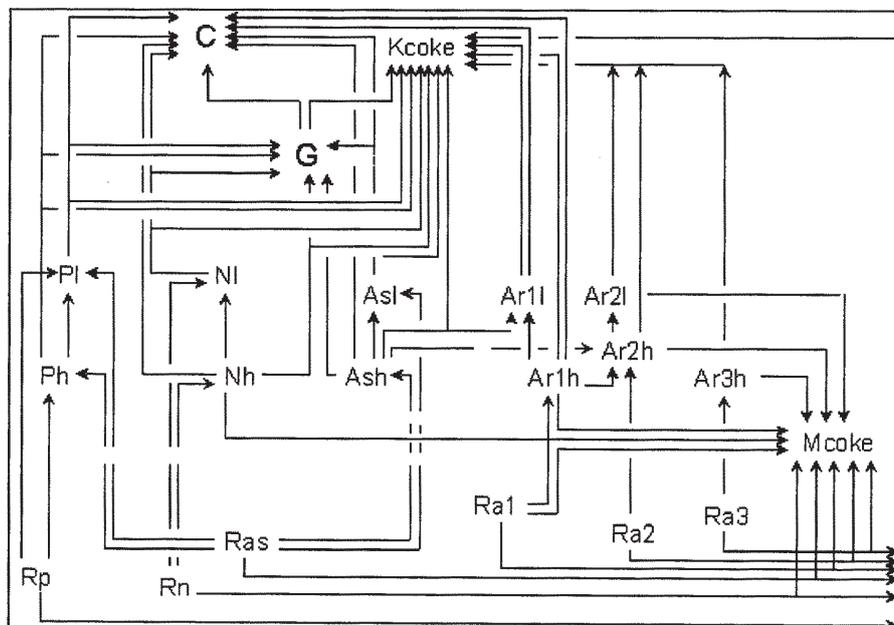
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Fig.1. Schematic for 21 lump reaction paths

No. Lump	Symbol	Description, NBP Range
1.	C	C lump – produces light gases
2.	G	Gasoline Lump C_5 < 430 °F (211 °C)
3.	Pl	Light Paraffins 430-650 °F (211-343 °C)
4.	Nl	Light Naphthenes
5.	Ar1l	Light 1-Ring Aromatics
6.	Ar2l	Light 2-Ring Aromatics
7.	Asl	Light Aromatic Ring Substituent Carbons
8.	Ph	Heavy Paraffins (650-950 °F) (343-510 °C)
9.	Nh	Heavy Naphthenes
10.	Ar1h	Heavy 1
11.	Ar2h	Heavy 2
12.	Ar3h	Heavy 3-Ring Aromatics
13.	Ash	Heavy Aromatic Ring Substituent Carbons
14.	Rp	Resid Paraffins (>950 °F) (510 °C)
15.	Rn	Resid Naphthenes
16.	Ra1	Resid 1-Ring Aromatics
17.	Ra2	Ra2 Resid 2-Ring Aromatics
18.	Ra3	Resid 3-Ring Aromatics
19.	Ras	Resid Aromatic Ring Substituent Carbons
20.	K coke	N/A Kinetic Coke
21.	M coke	Metals Coke

Table 1
TWENTY-ONE LUMP MODEL [16]

kinetics and are considered to occur in vapour phase. The kinetic expressions are integrated along the entire length of the riser, and are dependent on the catalyst bulk density, the coke on catalyst, MAT activity, the basic nitrogen, and metals content. The MAT activity and the basic nitrogen are provided by external model sources and affect the riser kinetics uniformly. The catalyst bulk density and the coke on catalyst are also integrated along the entire riser length, and are functions of pressure drop, coke formation, and molar expansion.

As mentioned above, reactor kinetics is based on the 21-lump kinetic model. The reaction pathways depicted in figure 1 represent paraffinic cracking, naphthenic ring opening, alkyl side chain cracking, ring condensation, kinetic coke formation from typical condensation reactions, and metals coke formation due to dehydrogenation. The reaction paths have been logically grouped to make yield parameterization more convenient. Thus, all the pathways leading to gas formation constitute one class, the gasoline pathways form other class, and so on. Following this approach, with only a small number of yield measurements during unit operation, the kinetic rate parameters for more than fifty reaction pathways can be easily tuned to match the yields of the industrial unit. To

match the specific product compositions observed in the industrial unit (provided that information is available), additional tuning of kinetic parameters for the reaction rates of paraffinic and aromatic hydrocarbons must also be performed.

This model divides the reactants and the reaction products into lumped aggregates of material classified by chemical nature and boiling point range. These lumps are similar to pseudo-components, but are also based on their molecular structure in addition to the boiling range for typical pseudo-component breakdowns. The molecular structures selected are based on likely reaction pathways and understood mechanisms known to exist in fluid catalytic cracking chemistry. The table below summarizes the lumps used in the Aspen model.

The riser – regenerator system is presented in figure 2.

The riser – reactor system is the most important part of the catalytic cracking unit as the cracking reactions take place in the riser. At the bottom of the riser, the gas oil feed comes in contact with the hot catalyst coming from the regenerator and, consequently, vaporizes. As a result, the cracking reactions start and the density of the feed decrease causing an increase in the velocity of the vapour phase. The increasing gas phase velocity accelerates the velocity of the catalyst and the riser behaves as a transport
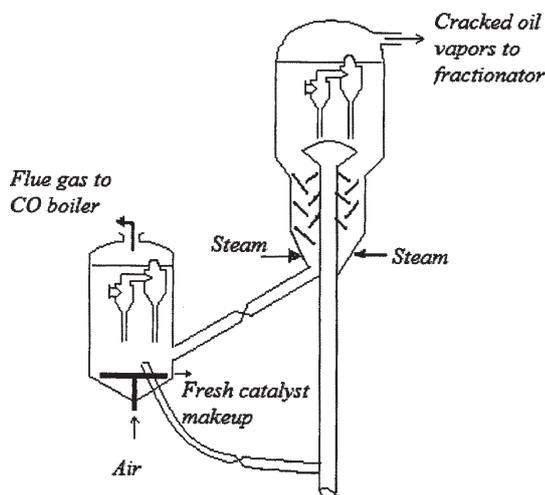
Fig. 2. Reactor-regeneration unit

bed reactor. The coke by product of the cracking reactions is deposited on the catalyst surface leading to a decreasing activity as the catalyst moves toward the exit of the riser. At the riser exit, the deactivated catalyst is separated from the vapour hydrocarbon products through specially designed riser termination devices and sent, through the striper, to the regenerator for burning off the coke from its surface and restore its catalytic activity. The vapour products are sent to the main fractionator for product separation and recovery [18].

The riser model consists of balance equations establishing the distribution of the reaction products, key point temperatures, pressure drop calculations, and coke distribution. The riser model is a segment of the fluidized riser that models the kinetics in the riser and includes the geometry of the riser for hydraulic and volume effects. It takes the hydrocarbon feed after the nozzle exit and combines it with the regenerated catalyst to take the material to the reactor. The chemistry in the risers is endothermic and uses the heat generated in the regenerator for the chemical transformations. This process is tracked along the length of the riser and is observed in the temperature profiles. In these profiles, the temperature of the hydrocarbon – catalyst mixture gradually drops from the entry zone to the riser exit into the reactor. These temperature drops are used in the models to determine catalyst flow rates [17].

The reactor model consists of three primary sub-models. As the hydrocarbon mixture enters the reactor vessel, a process of disengagement of the vapour hydrocarbons and catalyst particles begins. Cyclone models are the final stage of this disengagement at the top of the reactor. Material entering the cyclone models arrives there from the reactor free-board area. This area is represented by a model that sends material, primarily catalyst, to the dense bed of the reactor. From there the material enters the stripping zone where steam is used to remove as much of the remaining hydrocarbons as possible from the catalyst before it enters the spent catalyst transfer line [16].

Coke sources are separated into five distinct categories: kinetic coke, metals coke, coke resulting from the Conradson carbon of the feed, coke produced by non-vaporized feed, and coke formation controlled by the stripping performance. The Conradson carbon coke and non-vaporized feed coke are assumed to be physical types and are therefore deposited on the catalyst at the entrance of the riser, prior to any cracking or coking reactions. Kinetic coke and metals coke are both determined from kinetic expressions and are deposited on the catalyst gradually as

reactions proceed along the riser and in the reactor. The coke produced in the stripper is determined from the catalyst-to-oil ratio and the stripper performance curves.

The effluent from the riser splits into a solid part that proceeds to the stripping zone and a vapor phase, containing most of the reaction products and unreacted feed molecules, which heads up towards the cyclones. The solid part consists of the catalyst, the kinetic coke, and a fraction of vapour hydrocarbons adsorbed on the catalyst. Further cracking of the adsorbed hydrocarbons on the catalyst surface occurs in the dense bed while some of this material along with the stripping steam, proceeds to the cyclones where it mixes with vapour effluents from the riser.

Heat balances are performed at each mixing point of the above coupled system consisting of cyclones, dilute phase, and dense bed. These balances yield different temperatures for each point in the system: riser outlet (cyclone inlet), dense bed, and the reactor plenum chamber.

The reactor dense bed model is a differential-algebraic model taking into account a single catalytic cracking reaction for the low concentration of hydrocarbons adsorbed at the catalyst surface. It also performs the pressure drop calculation along the height of the catalyst bed, which can be set using pressure measurements in the industrial plant, or is to be specified directly in the model. In the latter case, the pressure drop is calculated.

The model for the reactor dilute phase (reactor plenum chamber) performs the mass balance, heat balance, pressure drop, reaction conversion and product distribution calculations. The dilute phase model represents the reaction volume that exists between the outlet of the riser and the inlet to the reactor cyclones [16].

Similar to the reactor model, the one for the regenerator consists of sub-models for the dense bed, the dilute phase, and the cyclones. In each of these sub-models heat balance, material balance and pressure drop calculations are performed. The regenerator dense bed kinetic model performs calculations for bubbling bed, heterogeneous combustion of coke, and both for the homogeneous and heterogeneous conversion of CO to CO_2. The regenerator cyclone model performs a two-phase, loading-based pressure drop calculation per cyclone. It completely returns the catalyst to the regenerator dense bed. This approach sets up a recycle of the catalyst that can alter the steady-state level of coke on the regenerated catalyst and the dense-bed temperature [16].

As discussed above, the user has access to a number of kinetic parameters of these models that can be adjusted as such that, by simulation, the model will reflect more accurately the behaviour of the industrial experimental unit. To this end, in this contribution we aim at using the Aspen HYSYS model and, by tuning the kinetic parameters of the model, to obtain in simulations results as close as possible to those observed in an industrial experimental plant. The resulting tuned model paves the way to optimize the functioning of the industrial plant with respect to product yields and operating parameters under different market conditions.

Experimental part

In the FCC model available in the HYSYS™ module of the ASPEN Engineering Suite package, the feedstock characteristics and the catalyst properties used in an industrial plant were introduced along with the geometrical characteristics of the equipment. The feed consists of a mixture of vacuum distillate and coker gas oil having the

Table 2
VACUUM DISTILLATE PROPERTIES

Property	Value
API Gravity	23.99
Specific Gravity 60F/60F	0.91
Distillation type	D1160
Initial Point (°C)	248.00
5 %	280.00
10 %	300.00
30 %	350.00
50 %	401.00
70 %	452.00
90 %	508.00
95 %	531.00
End Point	550.00
Sulfur content, %	1.80
Conradson Carbon Residue, %	0.26
Ramsbottom Carbon Residue,%	0.26
RI@20C Est. From Bulk Propr.	1.51
Ca Est, from Total Method	16.43

Table 3
COKER GAS OIL PROPERTIES

Property	Value
Top Cut Point, °C	345
Botttom Cut Point, °C	495
API Gravity	24.15
Specific Gravity, 60F/60F	0.9091
Basic Nitrogen Content, ppmwt	197.7
Total Nitrogen Content, ppmwt	593.00
Total/Basic Nitrogen Ratio, ppmwt	3.00
Sulf Content, %	2.27
Conradson Carbon Residue, %	0.01
Ramsbottom Carbon Residue,%	0.07
RI@20C Est, from Bulk Prop.	1.51
Ca Est, from Total Method	19.39
Nickel content, ppmwt	$4.71 \cdot 10^{-9}$
Distillation Data	
Initial Point (°C)	328.0
5 % (C)	338.8
10 % (C)	349.5
30 % (C)	383.9
50 % (C)	416.0
70 % (C)	446.9
90 % (C)	478.0
95 % (C)	488.1
End Point (C)	500.0

Table 4
PROPERTIES OF THE EQUILIBRIUM CATALYST USED IN THE EXPERIMENTAL INDUSTRIAL UNIT

Parameter	Value
V, ppm	297.9
Ni, ppm	344
Na, ppm	1556
Fe, ppm	8203
Cu, ppm	25.0
Catalyst inventory, kg	25000
Fresh catalyst adding, kg/h	19100
MAT, %	66.5
Kinetic Coke	0.87
Metals Coke	1.126
Feed Coke	1.237

characteristics presented in tables 2 and 3, while the properties of the equilibrium commercial catalyst used in the industrial unit are provided in table 4.

Subsequently, several sets of test data were recorded from the experimental industrial plant – in operation within a Romanian refinery – together with the operating parameters used during the test data collection. Data sets were collected from the industrial catalytic cracking plant for a period of 7 months using the same catalyst and the same feedstock. The test data sets collected cover feed characteristics, riser and regenerator temperatures, feed flow rate, riser and regenerator pressures, as well as main reaction products' yields (gasoline, diesel, coke, propylene, and gas).

Simulating the industrial plant using the actual geometrical characteristics, feed stock properties in Tables 2 and 3, catalyst characteristics in table 4 and the working riser temperature, the stripping steam pressure and temperature, regenerator bulk density, and the pressure regime, the model provides estimated values for different parameters, among which we have chosen the regenerator temperature and the gasoline, light cycled oil (LCO) and coke product yields to test the model adequacy to the industrial plant performance.

Using the values of the operating parameters collected from the experimental industrial unit during test data acquisition, simulations were run initially with the intrinsic values of the kinetic parameters of the model. In order to assess the adequacy of the model using the intrinsic values of the kinetic parameters, the values obtained by simulation for the yields of the main reaction products and for the regenerator temperature were plot in figures 3 - 6 against the experimental values collected during test data acquisition from the experimental industrial unit.

As shown in figures 3 to 6, simulation of the catalytic cracking unit using the intrinsic values of the kinetic parameters of the model leads to significantly different results compared to those obtained experimentally in the industrial plant. This indicates the need for model adjustment prior to obtaining an accurate simulation model that would be able to reproduce the functioning of the experimental industrial unit.

Results and discussions
Using the intrinsic values of model parameters, the deviation of the yields obtained by simulation were significant compared to the actual values obtained experimentally in the industrial plan. The most significant deviations were observed for the LCO and coke yields. In order to quantify the deviation of results obtained in simulation from their experimental values the comparison was made using as a benchmark the model accuracy percentage error, E, and root mean square deviation, σ, calculated with the following relations:

$$E = \left(\frac{|x_{calc} - x_{exp}|}{x_{exp}} \right) \cdot 100$$

$$\sigma = \left(\frac{\sum_{i=1}^{n} \left(x_{calc} - x_{exp} \right)^2_i}{n} \right)^{1/2}$$

Using the accuracy error, for the LCO yields calculated with the intrinsic values of the kinetic parameters of the model were as high as 35 %, with most of the experimental yields higher than those obtained in simulation. At the same time, the regenerator temperature values resulting from the simulation were 15 to 20° C higher than those observed experimentally. Because also the gasoline experimental yields have higher values than the calculated ones we conclude that a faster conversion of LCO to gasoline and of gasoline to gas is needed in order to improve the model adequacy.
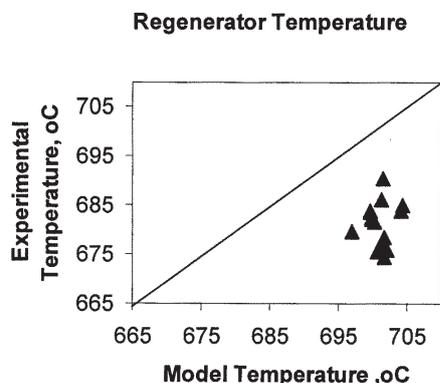
Regenerator Temperature

Fig. 3. Initial model response compared with real data on regenerator temperature

Gasoline Yield

Fig. 4. Initial model response compared with real data on gasoline yield

LCO

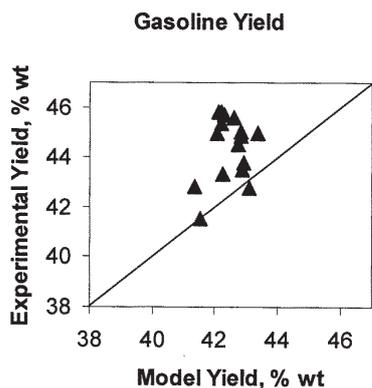Fig. 5. Initial model response compared with real data on LCO

Coke Yield

Fig. 6. Initial model response compared with real data on coke

The user has a number of kinetic parameters of the model that could be adjusted so that the model can give estimates that faithfully reflect the behaviour of the industrial plant. Because of the observations above we chose to modify several kinetic parameters. Their values were modified in successive steps and the simulation results were compared with test data from the industrial experimental unit.

We have started by adjusting the Basic Nitrogen Deactivation for G Pathways, Pre-exponential Factor for Gasoline Cracking, Ea/R for Gasoline Cracking, Distillate Naphthenic Cracking to G Lump, Distillate Aromatic Sidechain Cracking to G Lump, and the Distillate Aromatic Ring Cracking to G Lump. Adjusting the parameters of the Distillate Cracking to C lump group, leads to changes in regenerator temperature and gasoline yields. Even if the errors on the regenerator temperature slightly increases, this adjustment has a positive effect as it decreases the gasoline yield error from 4.77 to 0.21 %, which translates to a total error reduction by 95.6 %.

For the decrease of the coke yield error the Basic Nitrogen Deactivation for Coking Pathways, Fraction Concarbon to Coke, CRC Response Intercept, CRC Response Slope, Global Catalyst Deactivation Factor, and the Catalyst Activity / Coking Factor parameters were further adjusted. The CRC Response Intercept and the CRC Response Slope are two parameters that tune simulation responses to the presence of coke on the catalyst, primarily for partial burning units, while the Global Catalyst Deactivation Factor, and the Catalyst Activity / Coking Factor parameters in the MAT Tuning category adjust the effects of the equilibrium catalyst activity in simulation influencing yields and make-up rate [17]. Following several trials, by modifying the CRC Response and MAT Tuning group parameters, we succeeded to obtain much smaller values of the coke errors, the final values decreasing from an initial
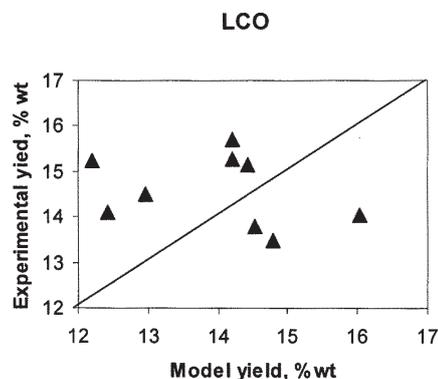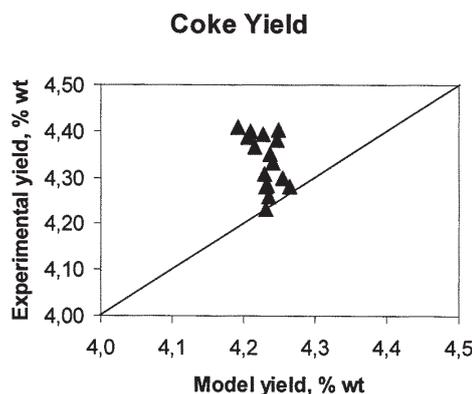
5.48 to 0.98%. Similarly, the LCO errors dropped from 13.47 to 1.64 %.

However, the Fraction Concarbon to Coke parameter is also important as a lower value causes an increase of the LCO and coke yield errors, but a better correlation for the gasoline yield and the regenerator temperature. The regenerator temperature is correlated with the amount of coke formed. Therefore, by reducing the regenerator temperature will eventually result in a lower coke yield, while maintaining a constant temperature in the reactor. Also, changing the flow of dilute phase (Regenerator Dilute Phase Flow Parameter) leads to values closer to those observed in the experimental industrial unit.

Regarding the LCO, it is known that getting a smaller yield of LCO is always associated with increased coke and gas yields. Thus, changing the values of the C group parameters, which are responsible for the formation of gases, adjusts the LCO yield values closer to the actual values obtained in the real catalytic cracking plant. The parameters in the C group are the Conversion of H2 to C2 Yields to Liquid Volume Basis, Distillate Paraffin Cracking to C Lump, Distillate Naphthenic Cracking to C Lump, Distillate Aromatic Sidechain Cracking to C Lump, and Light Gas Delumping to Hydrogen. Distillate paraffin cracking to C lump refers to the natural log of the pre-exponential factor for the first order cracking of light paraffin (L-P) to light gases (C-lump). Distillate naphthenic cracking to C lump represents the natural log of the pre-exponential factor for the first order cracking of light naphthenic (L-N) to light gases (C-lump). Distillate aromatic sidechain cracking to C lump is a parameter representing the natural log of the pre-exponential factor for the first order cracking of light aromatic sidechains (L-As) to light gases (C-lump). Finally, the C lump is used to calculate the light gases for methane through the pentanes, and is based on a correlation using the C lump produced in the kinetic paths and the composition of the feed. Within the kinetic system,

| Kinetic Parameter	Difference, %
Basic Nitrogen Deactivation	
Basic Nitrogen Deactivation for Coking Pathways	-191.38
Basic Nitrogen Deactivation for G Pathways	72.68
Basic Nitrogen Deactivation for Other Pathways	58.33
FOE for Gas Yields	
Conversion of H_2 to C_2 Yields to Liquid Volume Basis [MJ/m^3]	-0.006
Coke Distribution	
Fraction Concarbon to Coke	-1100
Gasoline Overcracking	
Pre-exponential Factor for Gasoline Cracking	1.08
Ea/R for Gasoline Cracking	-8.33
Regenerator Dilute Phase Flux Parameter [kJ/kg]	-8233.33
Distillate Paraffin Cracking to C Lump	-640.74
Distillate Naphthenic Cracking to C Lump	-1033.33
Distillate Aromatic Sidechain Cracking to C Lump	99.68
Distillate Naphthenic Cracking to G Lump	-60
Distillate Aromatic Sidechain Cracking to G Lump	-2.86
Distillate Aromatic Ring Cracking to G Lump	96
Methane from Feed Heavy Aromatic Sidechains	0.25
Entry zone EA over RT (1000 F)	0.01
Light Gas Delumping to Hydrogen	-0.002
CRC Response	
CRC Response Intercept	50
CRC Response Slope	73.68
MAT Tuning	
Global Catalyst Deactivation Factor	-29061.1
Catalyst Activity vs. Coking Factor	15.56

Regenerator Temperature
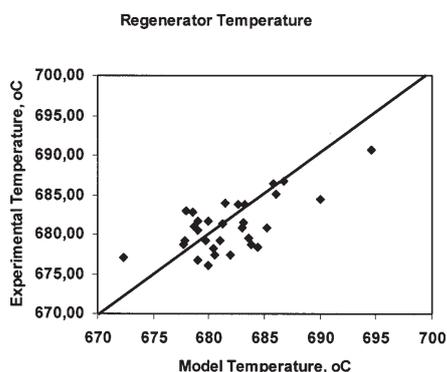
Fig. 7. Theoretical and experimental values correlation for the
regenerator temperature

Gasoline

Fig. 9. Theoretical and experimental values correlation
for gasoline

LCO

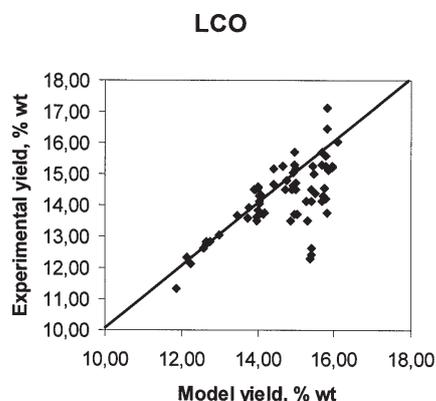Fig. 8. Theoretical and experimental values correlation for LCO
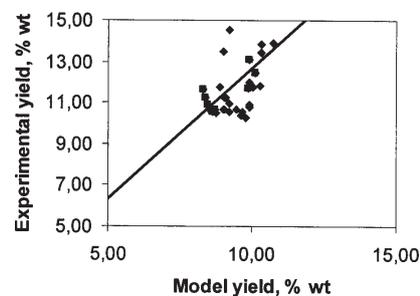
Bottoms

Fig. 10. Theoretical and experimental values correlation for
bottoms

the C lump component is divided directly into ten individual light-gas components and coke [17].

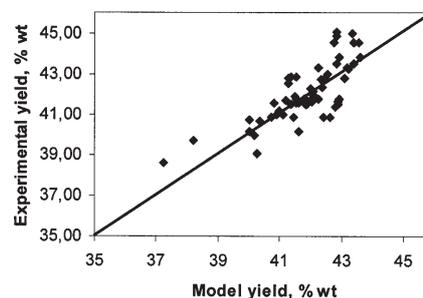The adjusted parameter values with errors less than 5 % and the smallest mean square deviations were chosen for the tuned model. Following successive attempts, the values of the kinetic parameters in table 5 were considered satisfactory, as simulating the functioning of the industrial unit using the conditions of the test data sets, the results predicted by the model accurately reflected the behaviour of industrial plant within a maximum error value of ± 4.88 %.
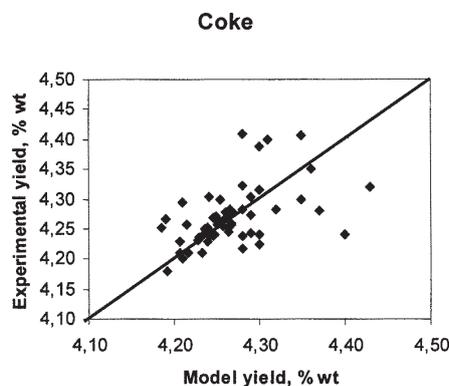
Fig. 11. Theoretical and experimental values correlation for coke

Fig. 12. Theoretical and experimental values correlation for diesel and bottoms

To illustrate the accuracy of the tuned model, values of the main variables of the process predicted by the model were plotted against the values of the same variables measured experimentally under the same conditions in the industrial plant. Figures 7-11 show the regenerator temperature and yields of gasoline, diesel and coke, respectively, as obtained from simulation and experimentally.

Simulations were performed using the model that gave the best response for gasoline, LCO and coke yields, and for the regenerator temperature, respectively. The model gave responses that remained in the parameters of minimum deviation for a longer period of time. Comparing simulation results obtained using this model with experimental results from the experimental industrial plant we obtained the results in figures 7-11.

A good correlation is observed with respect to the regenerator temperature and the yield of gasoline and coke. However, the LCO yield does not show a similar correlation. A possible explanation is that, in the data sets obtained from the refinery, there were not available information with respect to the amount of heavier distillates, while the simulation software provides data only in terms of LCO. Taking into account that for the feed we only had daily average compositions (constant ratio of vacuum distillate to coker gas oil is not always constant) and considering that heavy oil is included in the slurry residue, we considered more relevant a graphical representation of the sum of diesel and bottoms yields resulting from simulation versus the real data collected from the industrial unit. As expected, the data correlates significantly better, as it can be observed in figure 12. The calculated error is 4.36 %.

Using this model that accurately reflects the industrial plant operation it is possible to simulate different scenarios for industrial plant operation. A first attempt was to investigate the model response to changes in the feed flow rate and the reactor temperature compared to the real FCC unit data. Yields for the major reaction products, such as gasoline, diesel and coke were estimated through simulation for temperatures between 520 and 540 °C, and several feed flow rates.

Conclusions

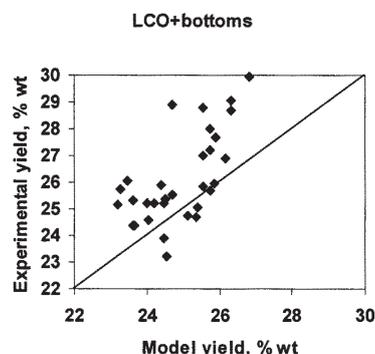We have successfully adjusted the intrinsic Aspen HYSYS fluid catalytic cracking model to accurately reflect the performance of an industrial catalytic cracking unit functioning in a Romanian refinery. The tuned model has been proven convergent for a wide range of operating parameters, and accurate within 4.88 % of overall error for the investigated range of parameters for which we had experimental values for comparison. The adjusted model can be further used to optimize the operation of the industrial plant.

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