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The Influence on Products Yield and Feedstock Conversion of Feedstock Injection Position along the Industrial Riser

Helver Crispiniano Alvarez-Castro[a], Victor Armellini[a], Milton Mori[*a], Waldir Pedro Martignoni[b], Raffaella Ocone[c]

[a] University of Campinas, School of Chemical Engineering 500 Albert Einstein Ave, 13083-970 Campinas, SP.
[b] PETROBRAS/AB-RE/TR/OT, 65 República do Chile Ave, 20031-912 Rio de Janeiro, RJ, Brazil.
[c] Chemical Engineering, Heriot-Watt university, Edinburgh EH144AS,UK.
mori@feq.unicamp.br

The fluid catalytic cracking (FCC) process is at the heart of a modern refinery oriented toward maximum gasoline and diesel production. Within the entire refinery process, this process offers the greatest potential for increasing profitability; even a small improvement in the gasoline yield it implies a substantial economical profit when dealing with a production of millions of barrels of gasoline a day. There are several articles published in the last two decades focusing the attention on 2-D and 3-D computational fluid dynamic models of the industrial riser of a circulating fluidized bed. Nevertheless, there are few research works published in the literature that include studies on how the localization of feedstock along the riser affects the yield products. A 3D hydrodynamic model coupled with a 12 lump kinetic model is presented in this work. Four different injection points in an FCC industrial riser were considered in order to evaluate the hydrodynamic behavior and their effect in the gas oil conversion and products yield. The equations were solved numerically by finite volume method using the Eulerian-Eulerian approach and a commercial CFD code, CFX version 14.0. Appropriate functions were implemented in the model via user defined functions considering the heterogeneous kinetics and catalyst deactivation. The results from the model were validated against the experimental industrial results and it was found that the conversion of gas oil and the production yield significantly change with the feedstock localization.

1. Introduction

Due to the large yield to products and the added value, fluid catalytic cracking (FCC) processes are central to the refinery operation. Within the entire refinery process, this process offers the greatest potential for increasing profitability; even with a small improvement in the gasoline yield it implies a substantial economic gain. The FCC process increases the H/C ratio by carbon rejection in a continuous process and it is used to convert the high boiling, high molecular, weight hydrocarbon fractions (typically a blend of heavy straight-run gas oil, light vacuum gas oil, and heavy vacuum gas oil) to more valuable products like gasoline, olefin gases, and other products (Barbosa et al., 2013).

The FCC process preheats the feedstock before it enters the riser; the feedstock is atomized by nozzles with the help of fluidization steam. The feedstock entering the riser contacts the regenerated catalyst. Cracking reactions initiate as soon as the feed is evaporated by the hot catalyst. An effective contact of the gasoil and the catalyst is essential for the desired cracking reactions. The increasing volume of the vapors acts as the means to carry the solids up the riser. Smaller gas oil droplets improve the availability at the catalyst active sites. Due to the high-activity of zeolite catalyst, the heat absorbed by the catalyst during regeneration provides the energy to evaporate and heats the feed to its desired reaction temperature (about 580 °C). The average heat of reaction, resulting from feedstock evaporation and the cracking reactions is endothermic. In most FCC risers, the feedstock injections are of an “elevated” type, meaning that they are located about 5–12

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m above the base of the riser. A very comprehensive description of this process is given by Sadeghbeigi (2012).
Computational fluid dynamic (CFD) is an essential tool for the design and performance evaluation of the fluid catalytic process and the process equipment in 3D as related by Lopes et al., (2011); for this reason an appropriated simulation of the phenomenological dynamics involved in the riser is very important to get more accurate predictions. The Eulerian-Eulerian approach has been used to represent phases where the solid is treated as a continuum for all fluid dynamic purposes (Nayak et al., 2005). To describe the reaction of the various components, the lumping approach has been used, where each lump is constituted by hundreds of different molecules in a specific range of molecular weight. The number of lumps may be increased to obtain a more detailed prediction of products distribution. The system is reduced to a finite number of lumps with each lump constituted by many components having similar characteristics and in a specific range of molecular weight (Ancheyta-Juárez et al., 1997). In the present work, a 12 lump kinetic model has been used to describe the catalytic cracking reactions in an industrial riser. This model presents the advantage of allowing for a better description of both products and feedstock, being one of the few complete models reported in the literature (Wu et al., 2009). The main objective of this work is to study the effect of feedstock positions along the riser in order to evaluate its effects on the conversion and product yields.

2. Mathematical Model

In the present work the governing equations and catalytic cracking kinetic models for the transient fluid model were taken and adapted from Alvarez-Castro et al., (2015). In order to study the heterogeneous kinetics and the particle phase deactivation, equations 8 to 11 were implemented in the CFX code. In Table 1 all equations are summarized.

Table 1 : Governing equations for transient two fluid models

Governing equation	Equation	N°
Continuity equations	$\frac{\partial}{\partial t}(\varepsilon_g \rho_g)+\nabla.\left(\varepsilon_g \rho_g u_g\right)=0$	(1)
	$\frac{\partial}{\partial t}(\varepsilon_s \rho_s)+\nabla.(\varepsilon_s \rho_s u_s)=0$	(2)
Momentum equations	$\frac{\partial}{\partial t}\left(\varepsilon_g \rho_g u_g\right)+\nabla.\left(\varepsilon_g \rho_g u_g u_g\right)=\nabla.\left[\varepsilon_g \mu_g \left(\nabla u_g+\left(\nabla u_g\right)^T\right)\right]+\varepsilon_g \rho_g g-\varepsilon_g \nabla p+M$	(3)
	$\frac{\partial}{\partial t}\left(\varepsilon_g \rho_g u_s\right)+\nabla.(\varepsilon_s \rho_s u_s u_s)=\nabla.\left[\varepsilon_s \mu_s \left(\nabla u_s+(\nabla u_s)^T\right)\right]+\varepsilon_s \rho_s g-\varepsilon_s G \nabla \varepsilon_s-M$	(4)
Turbulence: The k-epsilon mixture model		
Heat transfer model	$\frac{\partial}{\partial t}(\varepsilon_g \rho_g H_g)+\nabla.(\varepsilon_g \rho_g u_g H_g)=\nabla.(\varepsilon_g \lambda_g \nabla T_g)+\gamma(T_s-T_g)+\varepsilon_g \rho_g \sum_r \nabla H_r \frac{\delta C_r}{\delta t} -Q_R-Q_V$	(5)
	$\frac{\partial}{\partial t}(\varepsilon_s \rho_s H_s)+\nabla.(\varepsilon_s \rho_s u_s H_s)=\nabla.(\varepsilon_s \lambda_s \nabla T_s)+\gamma(T_g-T_s)$	(6)
Variation of the chemical species	$\frac{\partial}{\partial t}\left(\varepsilon_g \rho_g C_{g,l}\right)+\nabla.\left(\varepsilon_g \rho_g u_g C_{g,l}\right)=\nabla.\left(\varepsilon_g \Gamma_i \nabla C_{g,l}\right)+\hat{R}_l$	(7)
The rate equation for the generic reaction	$\hat{R}_{l,r}=-k_r.\rho_p.(\rho\alpha_i).\phi(t).F(N).F(A)$	(8)
Decay model based on coke content	$\Phi(t)=e^{(-\alpha t)}$	(9)
Alkaline nitrides	$F(N)=\frac{1}{1+k_N C_N t_C/F_{c/o}}$	(10)
Polycyclic aromatic adsorption	$F(A)=\frac{1}{1+k_A(C_A+C_R)}$	(11)
Arrhenius' equation	$k_r = k_r^0 \exp\left(\frac{E_r}{RT}\right)$	(12)
Arrhenius equation (any temperature and dependent on the hold)	$k_c(T,\varepsilon_s)=k_{r,550°C}(\rho,\varepsilon_s)\exp\left[-\frac{E_r}{R}\left(\frac{1}{T}-\frac{1}{550°C}\right)\right]$	(13)

3. Simulation

A study of the influence of the height of the feedstock injection was simulated for the conditions shown in Table 2.

Table 2: Operation conditions varying the height of feedstock injection

ITEM	CASE A VALUE	CASE B VALUE	CASE C VALUE	CASE D VALUE
Reaction temperature (K)	793.15	793.15	793.15	793.15
Fluidization steam (%)	3	3	3	3
Flux of fresh feedstock (t/h)	124.46	124.46	124.46	124.46
Inlet temperature of fresh feedstock (K)	543.15	543.15	543.15	543.15
Catalyst temperature at riser inlet (K)	913.15	913.15	913.15	913.15
Ratio of catalyst to oil	8.1	8.1	8.1	8.1
Height of feedstock injection (m)	5	15	25	35

The finite volume method technique, using the commercial software ANSYS CFX 14.0 as a tool to discretize and solve the governing equations, was implemented. First, an industrial riser (wye section) with the specifications was considered; the height of feedstock injection, as shown in Figure 1, was varied; then, the geometries considered were meshed. A hybrid mesh with 800 thousand control elements was built and applied in this work; meshes between 700 to 900 thousand control elements are independent and recommended for a good representation of industrial risers as showed in previous works by Alvarez-Castro et al., (2012). Gasoil is fed totally vaporized due to the high-efficient nozzles employed by KBR-technology (2009). 400 kJ/kg was adopted in the simulation as the heat needed for the evaporation of the liquid droplets taken from Alvarez-Castro et al., (in press). A time step 10-3 s and monitoring Courant number of less than one to guarantee good results were assumed. About twelve days of running were necessary to predict a period of time (15 s) long enough to show that the variables have a cyclic behavior.

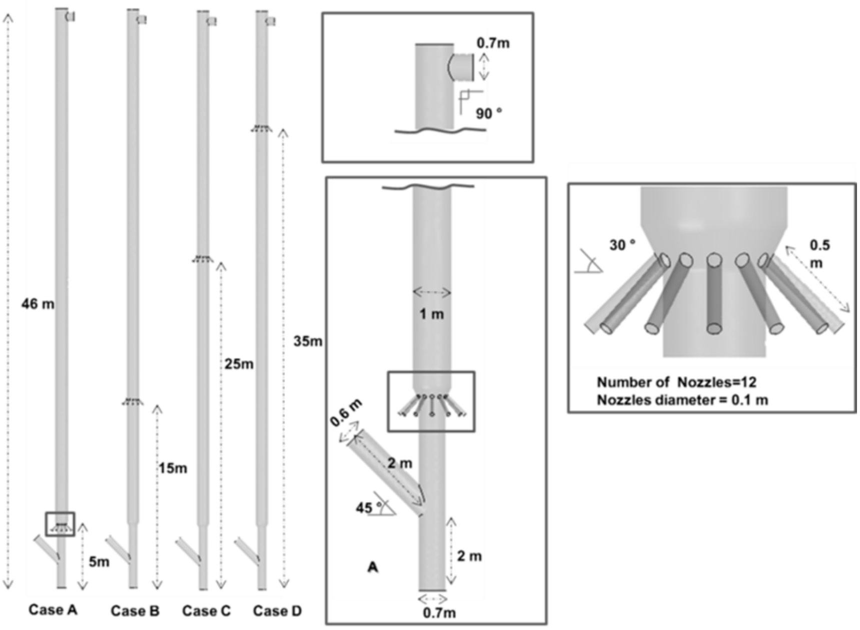

Figure 1 : Riser geometry varying the height of feedstock injection

4. Results and Discussion

4.1 Comparison of fluid dynamic profiles for the height of feedstock injection

Figure 2 shows the volume fraction profiles; cases A and B show profiles with greater homogeneity of the catalyst distribution over the riser, while the profiles of the cases C and D show little or no catalyst after the feedstock injections. In all cases the fluidization velocity is the same but due to the reagent feed and the catalytic cracking reaction, there is an increase in the gas velocity so that the catalyst has a larger displacement along the riser; cases C and D failed to reach this point.
In Figure 3, the global temperature behavior in the riser can be observed. Cases A and B have a higher overall temperature profile in the last meters of the riser while cases C and D show a low temperature in the last meters of the riser height; this behavior is mainly due to two situations: 1) the endothermic character of the reaction and 2) the low or zero fraction of catalyst in the last meters of the riser height in cases C and D.

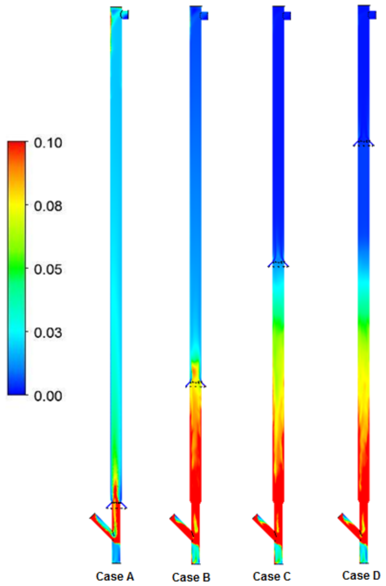

Figure 2 : Catalyst volume fraction profiles in axial planes

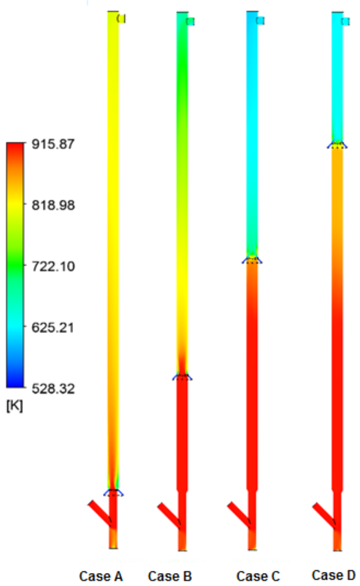

Figure 3 : Temperature contour profiles in axial plane

4.2 Comparison of the kinetic performance for the heights of feedstock injection.

A comparison of the conversion and yield for all the cases studied can be seen in Figure 4. It is important to show, as a first result, that the model simulation Case A was validated against industrial data from Wu et al., (2009) for the same conditions simulated. It can be also observe that the cases A and B show a yield of gasoline and diesel according to the normal converting process, while cases C and D show a very low level of conversion, thereby demonstrating that the heights recommended for the feedstock injection should not be greater than 15 m. In general, the effect of feeding at different riser heights (above 5 m) causes an effect on the residence time and the distribution of the temperature profile which leads to behavior similar to the case when the fluidization velocity is too high that produces a low conversion of feedstock.

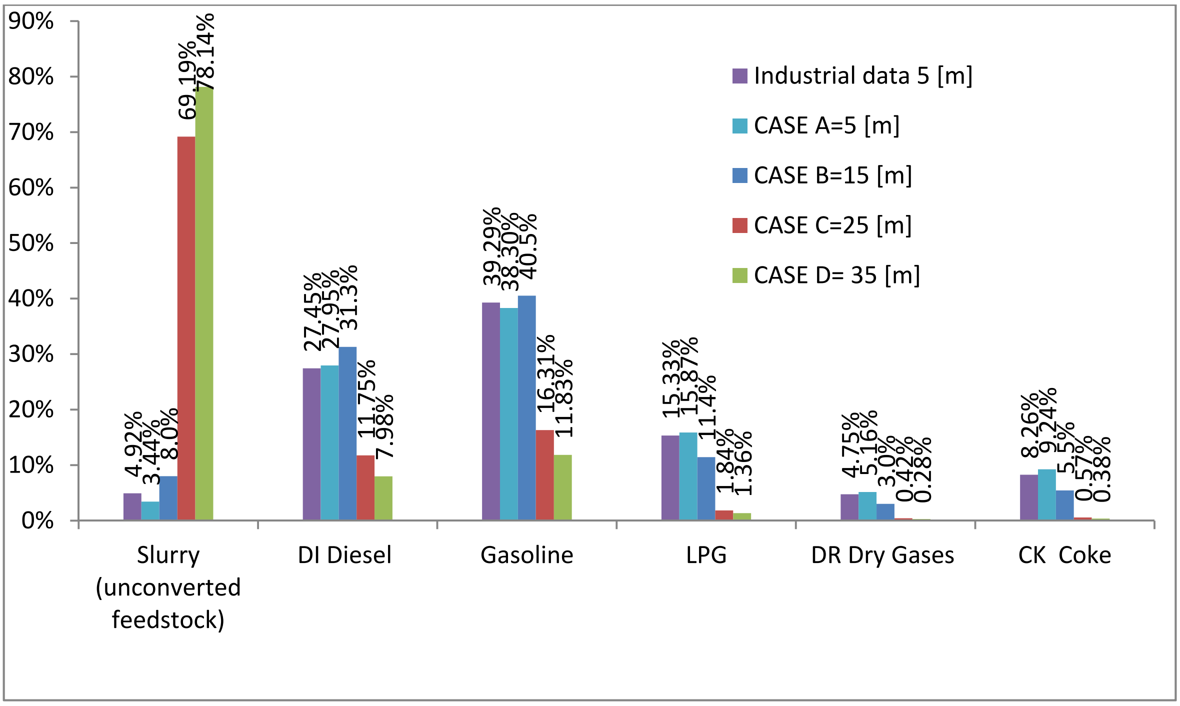

Figure 4 : Comparison of the product yield for different height of feedstock injection

5. Conclusions

By comparing the simulation results with the experimental data, it can be concluded that the model well predicts the real process; therefore, the model can be employed as a tool helping the design and operation tests in industrial FCC risers of FCC.
A systematic investigation has been carried out to study the influence of the feedstock injection heights on the riser performance demonstrating their relevant role in the process. The recommend heights for the feedstock injection was also identified and those coincide with Cases A and B.

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