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Effect of Injection Zone on Catalyst and Gas Homogenization in FCC Riser

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In the fluid catalytic cracking (FCC) process, the riser is the most important equipment where the reaction taken place. In the riser operation, the feedstock is fed through injectors and mixed with catalyst and steam. A good design and localization of the injectors to ensure rapid evaporation of the feedstock and a good contact of the droplets with the catalyst is important to improve the process efficiency due to a better reagent distribution which ensures a good feedstock conversion and product yields. Hydrodynamic modelling, heat transfer and cracking reactions were studied in this paper using Computational Fluid Dynamic (CFD) in order to evaluate the effect of nozzles design and configuration on the homogeneity of the gas-solid distribution. A 3D model was solved with the Eulerian – Eulerian approach using ANSYS/CFX version 14.0 as calculation tool. The simulation results showed that the distribution of gas-solid depends significantly on the configuration of the feedstock injectors.

1. Introduction

The fluid catalytic cracking unit (FCCU) converts residues of oil gas vacuum (VGO) of low commercial value into a number of derivatives like gasoline , diesel , LPG, dry gases and coke with high economic value; it is considered by many authors (e.g., Fahim et al., 2010) one of the most important and profitable processes in petroleum refineries. The FCCU can be divided into three main sections: riser, separator and regenerator. The catalytic cracking occurs mainly in the riser, the riser of FCCU is a long vertical tube with a high ratio height/diameter, in which the catalyst and gasoil enter in contact and the reaction takes place. The bottom region of the riser is known as the lift region; the catalyst is fed at the bottom of the lift region and then it is fluidized by steam which is fed at the bottom of the riser. The feedstock is injected into the riser through nozzles, which aim to atomize the gasoil in small droplets and introduce them into the riser increasing the contact area between the catalyst and the gasoil. Additionally, the nozzles assure appropriate feed distribution inside the riser to minimize zones of high concentration of catalyst thus avoiding high temperature gradients that are not beneficial to the process.
The injection zone is considered the most complex region of the reactor, due to the intense turbulence and not homogeneous regime, resulting in high gradients of temperature and concentration. According to previous works (e.g., Mauleon and Coorcelle (1985), the nozzles' design has a big influence on the hydrodynamics and catalytic cracking, thereby influencing the yield.
Theologos et al. (1997) presented a one-dimensional model with 10-lumps kinetic model to describe the catalytic cracking reactions and studied the influence of nozzles' number on the reactor performance. The simulation result showed that the yield of the desired reaction is improved by increasing the nozzles' number, since it provides a more homogeneous catalyst distribution. Lopes et al. (2011a) used a three-dimensional model for gas-solid flow using a 4-lumps kinetic model. The results also showed that non-uniform catalyst distribution affects the reactor performance. Furthermore, the referenced authors emphasized that the use of

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more real injection designs generate a phase distribution which approximates better the real process. Li et al. (2013) used a 14-lump model to simulate the FCC riser. The simulation result showed that the feedstock injection velocity and the angle of injection influence the products yields, while the injector position did not show significant influence. Wolschlag et al. (2010) compared a conventional nozzle configuration with an arrangement of dual radial distribution; the results showed that the injector’ arrangement affects the catalyst distribution.

Since the injection of gas oil through nozzles affects significantly the riser performance, and consequently the economic performance of the FCCU, the purpose of this study is to predict the hydrodynamic behaviour in an industrial riser, using a CFD as a tool to evaluate different nozzle arrangements, in order to estimate the effect in the homogenization with the aim improving the catalyst distribution in the riser.

2. Mathematical model

A 3-D model was used to describe the flow transport phenomena with an Eulerian-Eulerian approach. In this approach the conservation equations of mass, momentum and energy (1-6) in Table 1, are solved simultaneously. To calculate the Nusselt number, Equation 11 is used, according with the Ranz-Marshall (Lopes et al., 2011b). As the flow regime in the FCC is considered to be turbulent, the K-ε model is used. The kinetic 12 lump model introduced by Wu (2009) is chosen to represent the reactions.

Governing equations

Continuity equations of gas and solid phases

$$\frac{\partial}{\partial t}\left(\varepsilon_g\rho_g\right)+\nabla.\left(\varepsilon_g\rho_g u_g\right)=0 \tag{1}$$

$$\frac{\partial}{\partial t}\left(\varepsilon_s\rho_s\right)+\nabla.\left(\varepsilon_s\rho_s u_s\right)=0 \tag{2}$$

Momentum equations of gas and solid phases

$$\frac{\partial}{\partial t}\left(\varepsilon_g\rho_g u_g\right)+\nabla\cdot\left(\varepsilon_g\rho_g u_g u_g\right)=\nabla.\left[\varepsilon_g\mu_g\left(\nabla u_g+\left(\nabla u_g\right)^T\right)\right]-\varepsilon_g\nabla p+\varepsilon_g\rho_g g+\beta\left(u_s-u_g\right) \tag{3}$$

$$\frac{\partial}{\partial t}\left(\varepsilon_s\rho_s u_s\right)+\nabla\cdot\left(\varepsilon_s\rho_s u_s u_s\right)=\nabla.\left[\varepsilon_s\mu_s\left(\nabla u_s+\left(\nabla u_s\right)^T\right)\right]-\varepsilon_s \mathrm{G}\nabla\varepsilon_s+\varepsilon_s\rho_s g+\beta\left(u_g-u_s\right) \tag{4}$$

Energy equation of gas and solid phases

$$\frac{\partial}{\partial t}\left(\varepsilon_g\rho_g H_g\right)+\nabla\cdot\left(\varepsilon_g\rho_g u_g H_g\right)=\nabla.\left[\varepsilon_g\lambda_g\nabla T_g\right]+h_{gs}\cdot\left(T_s-T_g\right)+\varepsilon_g\rho_g\cdot\sum_r\Delta H_r\cdot\frac{\partial C_r}{\partial t} \tag{5}$$

$$\frac{\partial}{\partial t}\left(\varepsilon_s\rho_s H_s\right)+\nabla\cdot\left(\varepsilon_s\rho_s u_s H_s\right)=\nabla.\left[\varepsilon_s\lambda_s\nabla T_s\right]+h_{gs}\cdot A_{g/s}\left(T_g-T_s\right) \tag{6}$$

Supplementary models and correlations

Drag force (Gidaspow 1994)

$$\beta=\begin{cases}150\dfrac{\varepsilon_s^2\mu_g}{\varepsilon_g d_s^2}+\dfrac{7}{4}\dfrac{\left|u_s-u_g\right|\varepsilon_s\rho_g}{d_s} & \varepsilon_s>0{,}2\\[2ex] \dfrac{3}{4}C_D\dfrac{\left|u_s-u_g\right|\varepsilon_s\varepsilon_g\rho_g\varepsilon_g^{-2{,}65}}{d_s} & \varepsilon_s<0{,}2\end{cases} \tag{7}$$

$$C_D=\begin{cases}0{,}44 & Re_s>1000\\ \dfrac{24}{Re_s}\left[1+0{,}15\left(Re_s\right)^{0{,}687}\right] & Re_s<1000\end{cases} \tag{8}$$

Solid pressure (Gidaspow 1994)

$$G=G_0 exp\left[c\left(\varepsilon_s-\varepsilon_{s,max}\right)\right] \tag{9}$$

Interphase heat transfer coefficient

$$h_{gs}=\frac{\lambda_g.Nu}{d_p} \tag{10}$$

Ranz-Marshall correlation

$$Nu=2{,}0+0{,}6.Re^{0{,}5}.Pr^{0{,}33} \tag{11}$$

3. Simulation

The geometry is adapted from Alvarez-Castro et al (2014). The reactor has three independent entrances: a steam fluidization inlet in the bottom, a lateral inlet for the catalyst on the bottom side and a feedstock inlet composed by eight nozzles with an angle of 45° relative 5 m above of the bottom. The geometry, with a detailed sketch of the nozzles is shown in Figure 1. The computational mesh is composed of approximately 1,300,000 control elements, the mesh are in agreement with previous works of Alvarez-Castro et al. (2012).

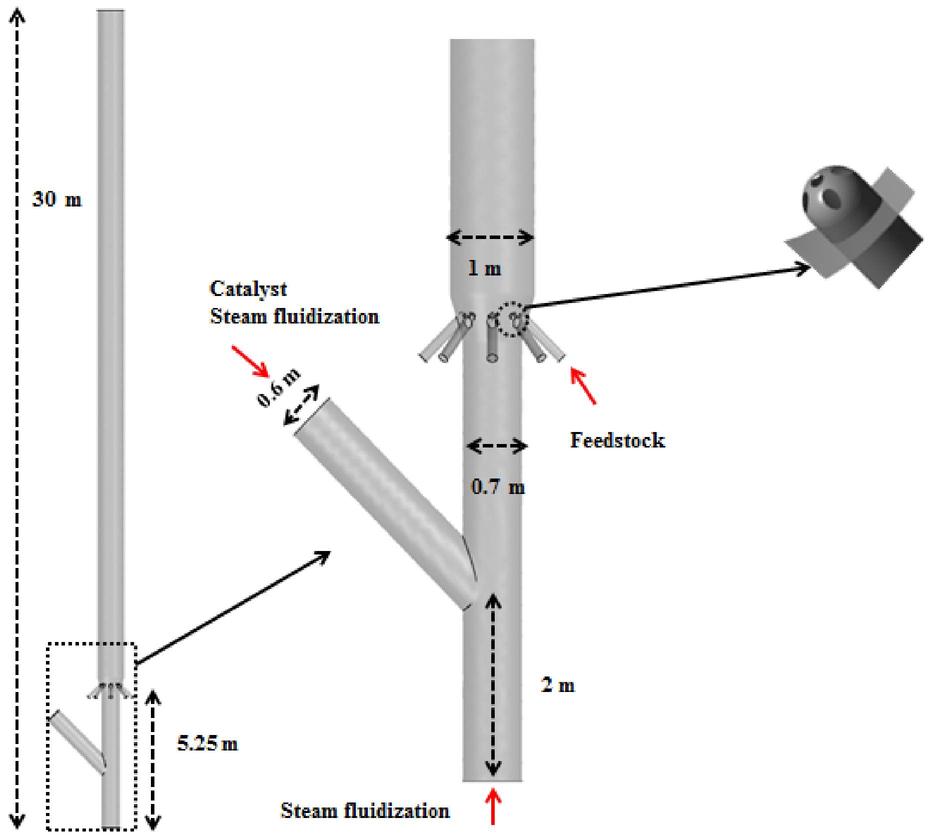

Figure 1: Riser Geometry

The operating conditions used in this work were taken from Chang et al. (2012), as shown in Table 2. No slip boundary conditions were used at the wall for the gas phase and free slip conditions for solid phase. A high resolution interpolation scheme as interpolation method and a convergence criterion RMS (root mean square) of 10^{-4} with a time step of a 10^{-3} were used. The simulation time was 15 s.

Table 1: Operating conditions (Chang et al., 2012)

Reaction temperature (K)	793.15
Pressure drop (kPa)	163
Reaction time (s)	3.22
Flux of fresh feedstock (t/h)	124.46
Inlet temperature of fresh feedstock (K)	543.15
Catalyst temperature at riser inlet (K)	913.15
Ratio of catalyst to oil	8.1

It is known that non-uniform catalyst distribution generates region with high temperatures, thus favouring the thermal cracking reaction. To improve the mixing between feedstock and catalyst three different feedstock nozzle arrangements were proposed, as show in Figure 2. In Case 1, the nozzles are evenly distributed around the riser at the same height; in Case 2 the nozzles are distributed interchangeably relative to the vertical axis; and in Case 3 the nozzles are distributed in pairs around the reactor.

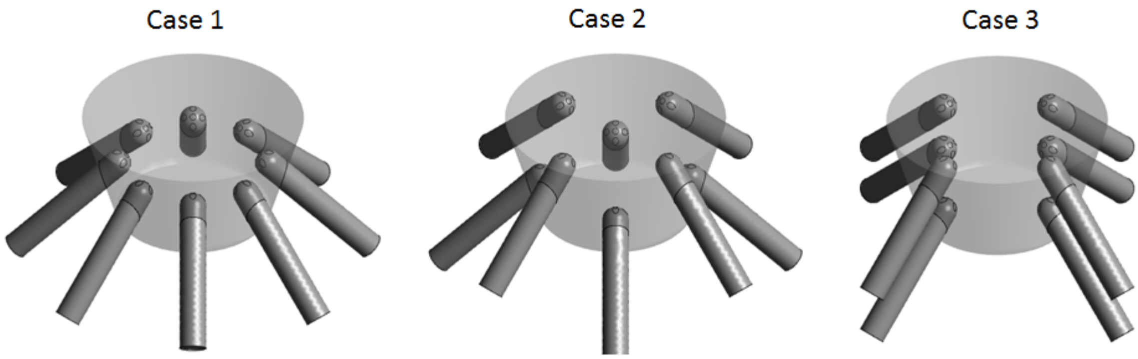

Figure 2: Feedstock nozzles arrangement proposed

4. Results

Figure 3 shows the catalyst volume fraction in cross-section at different heights. It can be observed that, after feedstock injection zone, (height 5.4m), the catalyst is not uniformly distributed for all cases. As the riser height increases, the flow becomes more diluted and an increase in catalyst velocity is observed: this feature improves the homogeneity of the flow. A different fluid dynamic profile along the riser height for the three cases was observed and it can be concluded that, the nozzles arrangement has significant influence on the catalyst volume fraction distribution.

For all cases, it can be observed that in the injection zone the catalyst particles are randomly distributed; this feature differs from related previous works in the same area (Alvarez-Castro et al (2015)), which that reported the catalyst particles are shift toward riser center due to push force of the gas oil injected. Also, it can be noted that the injection design and configuration have a big influence on the fluid dynamic profile.

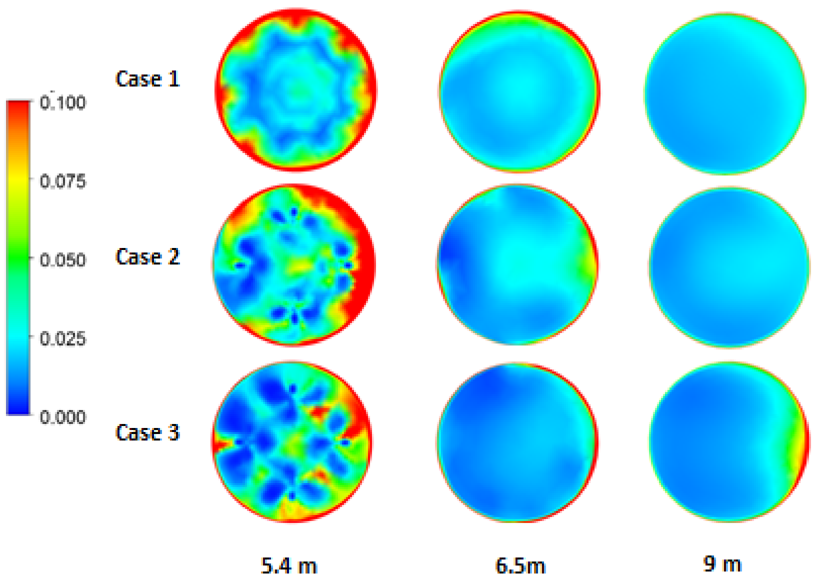

Figure 3: Catalyst volume fraction in cross-sectional planes

In order to establish the better nozzles arrangement, the histogram of the mean catalyst volume fraction in the planes at 6 m and 9 m height, is presented, as show in Figures 4 and 5, respectively: the dashed line represents the average of the catalyst distribution in the cross-sectional area at the same height in each case and the y axis indicates the amount of catalyst in the radial direction. It can be seen, in the Figure 4 (6 m), that all cases show a similar profile, with the higher catalyst concentration at the wall and a diluted central region; Case 1 presents a lower catalyst volume fraction at the wall in relation the other cases. At 9 m height a greater accumulation of catalyst at the wall is observed: however the difference between the amount of catalyst at the wall and at the centre decreases. Case 1 presents again the best homogenization.

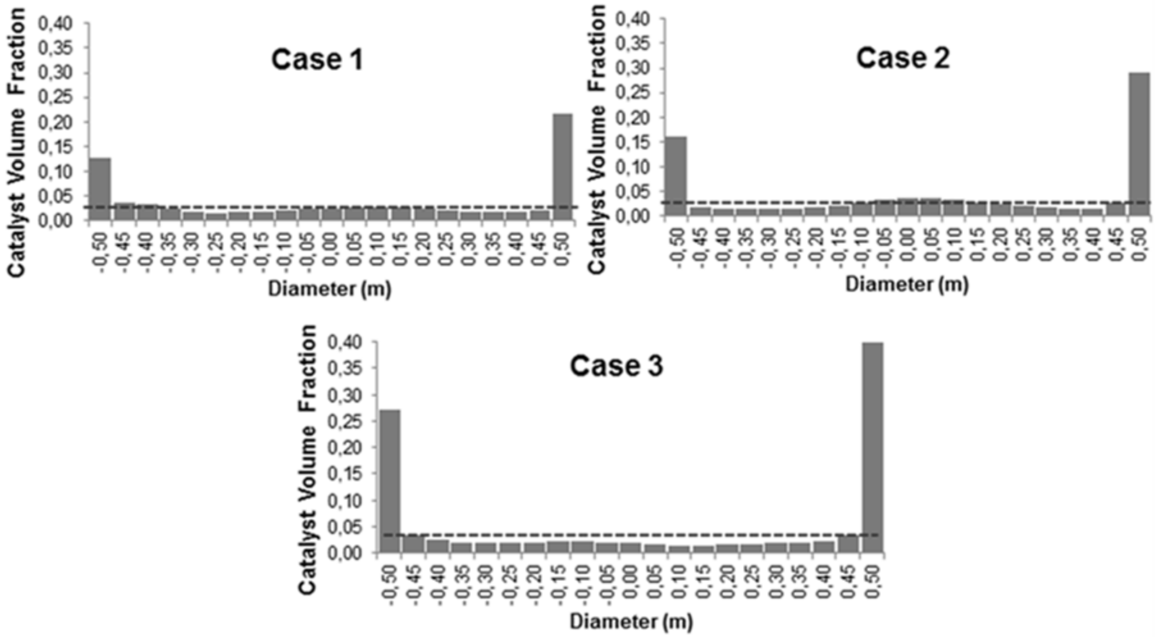

Figure 4: Radial distribution of catalyst volume fraction at 6 m

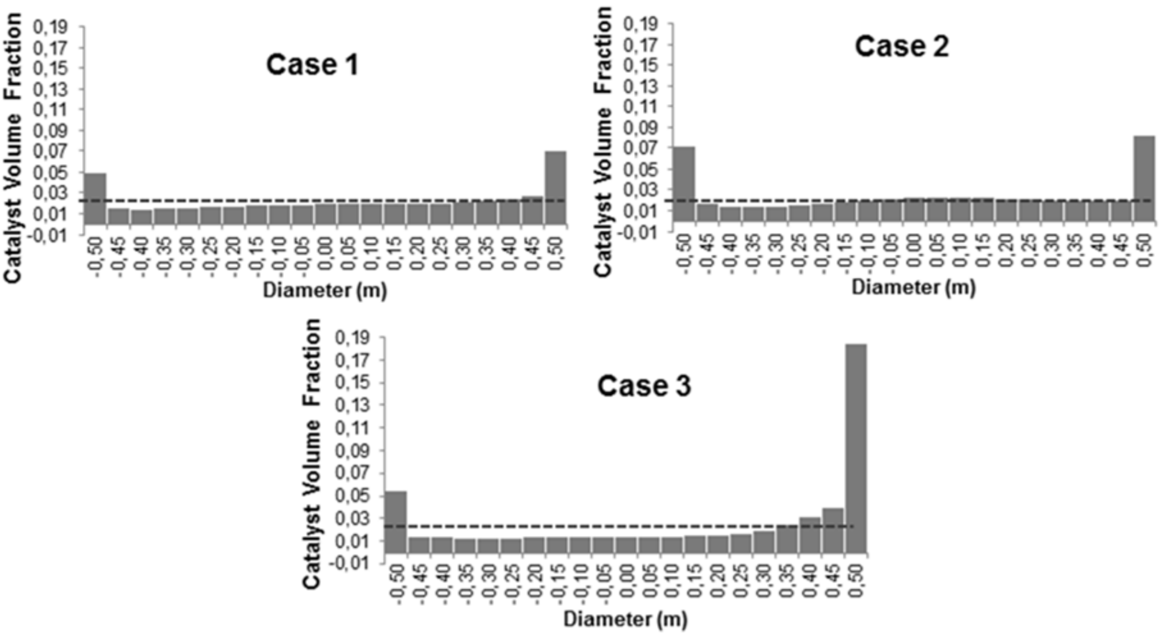

Figure 5: Radial distribution of catalyst volume fraction at 9 m

5. Conclusions

The results show a good solid distribution profile which improves the homogenization and minimizing the core annulus effect. The arrangement of the nozzles has significant influence on the hydrodynamics and mixing between the gas and solid phase improves. The design of nozzle used resulted in a different fluid dynamic profile found in previous works, providing a more homogeneous distribution catalyst immediately after the injection zone. Thus the simulation results show that it is possible to improve the catalyst distribution in the riser, through the use appropriate arrangements and nozzle design.

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