



Original Research Paper

The influence of the fluidization velocities on products yield and catalyst residence time in industrial risers

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ABSTRACT

The hydrodynamic characteristic of the industrial riser used in the fluid catalytic cracking (FCC) process has been simulated. A gas–solid flow model was developed which describes a 3D industrial set-up. By combining the hydrodynamics with a reaction model the yields of the different product families were obtained with good precision. To represent the kinetic behavior, a twelve-lump model with catalyst deactivation was adopted to represent the kinetic behavior. A tracer technique for catalyst residence time, corresponding to different fluidization velocities, was also considered. The Eulerian–Eulerian approach was adopted and solved by ANSYS CFX 14.0. The results show predictions for fluidization velocities and residence time which should be adopted to get better product yields in the industrial process. The results are compared with data taken in an industrial plant. The model furnishes valuable information on the impact of the riser hydrodynamics on the product quality.

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1. Introduction

The description of the conservation of mass and energy in gas–solid fluidization is an inherently complicated task, especially due to the lack of detailed knowledge of the nature of the interaction between phases. The phenomena that occur in this kind of systems pose challenges both in terms of mathematical modeling and appropriate numerical methodology to be implemented; consequently, a large number of publications in the field have flourished in the last decades. An application of gas–solid fluidization is the fluidized catalytic cracking (FCC) process which is the most important conversion process in petroleum refineries. In this process, a feed stream, consisting mainly of heavy distillates or residues, is broken into small and more valuable molecules using a solid catalyst. As far as computational resources are available, two approaches are used to predict the behavior of the solid and gas phases in this reacting gas–solid fluidization process, namely the Eulerian–Eulerian and the Eulerian–Lagrangian methods. In the Eulerian–Eulerian methodology both phases are interpreted as a continuous medium and the solid and gas are linked through solid–gas

interaction terms. The Eulerian–Lagrangian methodology assumes that the gas phase is a continuum whilst the solid is considered in terms of discrete particles; the phase coupling is considered by using semi-empirical terms. The Eulerian–Lagrangian approach has been largely used to describe the behavior of particulates in solid–gas flows; however such method requires a large computational effort [1]. In the present work, the Eulerian–Eulerian approach has been used due to less computational effort required; this approach has been followed by a number of the research groups [2–4]. In the Eulerian–Eulerian methodology the solid phase is treated as a continuum for all fluid dynamic purposes; the reactive system is then studied through the lumping approach which is shown to be very powerful when a large number of components is involved [5–9]. 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 [5].

In the present work, a 12 lump kinetic model was implemented to describe catalytic cracking reactions. 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 literature by Wu et al. [10]. The 12 lump model is considered in conjunction with the hydrodynamic model of the industrial riser of the FCC unit; the full model (including hydrodynamics and reactions) is thus solved. In addition, a computational fluid dynamic

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Nomenclature

C_i	molar concentration of component I (k mol m^{-3})	$Q(t)$	mass flux of tracer at the inlet of the secondary inlet (kg/s)
C_d	drag coefficient (–)	t_m	mean residence time (s)
C_G	constant of elasticity modulus function (Pa)	t_{min}	minimum residence time (s)
C_μ	constant 0.09	t_{max}	maximum residence time (s)
$C_{\epsilon,1}$	constant 1.44		
$C_{\epsilon,2}$	constant 1.92		
d	particle diameter (m)		
E	activation energy (J mol^{-1})	Greek letters	
g	gravitational acceleration ($\text{m}^2 \text{s}^{-1}$)	M	interphase momentum transfer ($\text{kg m}^{-3} \text{s}^{-1}$)
G	elasticity modulus (Pa)	ϵ	volume fraction (–)
H	static enthalpy (J mol^{-1})	\in	turbulence dissipation rate ($\text{m}^2 \text{s}^{-3}$)
k	kinetic constant of reaction or turbulent kinetic energy ($\text{m}^3 \text{k mol}^{-1} \text{s}^{-1}$) ($\text{m}^2 \text{s}^{-2}$)	\emptyset	catalyst decay function (–)
k^0	pre-exponential factor ($\text{m}^3 \text{kmol}^{-1} \text{s}^{-1}$)	γ	interphase heat transfer coefficient ($\text{Wm}^{-2} \text{K}^{-1}$)
k_c	deactivation constant ($\text{kg}_{cat} \text{k mol}^{-1}$)	Γ	diffusivity ($\text{kg m}^{-1} \text{s}^{-1}$)
Nu	Nusselt number (–)	λ	thermal conductivity ($\text{Wm}^{-1} \text{K}^{-1}$)
p	static pressure (Pa)	μ	molecular viscosity (Pa s)
p^k	shear production of turbulence (Pa s^{-1})	ρ	density (kg m^{-3})
Pr	Prandtl number (–)	σ_k	constant 1.00
q_i	specific coke concentration ($\text{k}_{mol} \text{kg}_{cat}^{-1}$)	σ_ϵ	Constant 3.00
R	reaction rate or universal gas constant ($\text{k}_{mol} \text{m}^{-3} \text{s}^{-1}$) ($\text{J mol}^{-1} \text{K}^{-1}$)	$C_{\epsilon,1}$	Constant 3.00
Re	Reynolds number (–)	$C_{\epsilon,2}$	Constant 3.00
T	static temperature (K)	φ	tracer concentration (kg/m^3)
\mathbf{u}	velocity vector (ms^{-1})		
Q_R	heat of cracking reactions (J Kg^{-1})	subscripts	
Q_V	energy lost in gasoil vaporization (J Kg^{-1})	g	gas phase
$A(t)$	accumulated residence time distribution function (–)	s	solid phase
D_φ	diffusion coefficient of tracer (m^2/s)	R	reaction
$E(t)$	residence time distribution function (–)	lam	Laminar
		$turb$	turbulent

(CFD) evaluation is undertaken to evaluate the residence time, the velocity profiles and their effect on the conversion and products yield.

2. Process description (Feed Nozzles–Riser)

The conversion section is the core of the FCC unit. The riser is the reactor. Effective and homogeneous contact of the reagents and the catalyst is vital for the desired cracking reactions. The feedstock is atomized by nozzles with the help of steam. Smaller gasoil droplets improve the availability at the catalyst active sites. Due to the high-activity of zeolite catalyst, practically all of the cracking reactions take place between 1 and 3 s. Generally, the feed injections are situated about 15–40 ft (5–12 m) above the base of the riser. Depending on the FCC design and on the riser diameter, the number of feed nozzles can vary from 1 to 15. Cracking reactions initiate as soon as the feed is evaporated by the hot catalyst. The increasing volume of the vapors acts as the means to carry the solids up the riser. The hot catalyst provides the required heat to vaporize the reagent and bring its temperature to the required cracking temperature, compensating, at the same time, for the decreasing in temperature due to the endothermic heat of reaction. According to the feed pre-heat, the ratio of catalyst to oil is normally in the range of 4:1 to 10:1 by weight. The catalyst temperature ranges between 1250°F and 1350°F (677–732 °C). The reactor temperature is often in the range of 925–1050°F (496–565 °C). Typical risers are planned for an outlet gas velocity of 40–60 ft/s (12–18 m/s). The average hydrocarbon and catalyst residence times are about 2 and 10 s depending on the feedstock quality and operating conditions. A hydrogen-deficient “coke” is deposited on the catalyst, decreasing catalyst activity and thus representing a concern for the efficiency of the cracking reactions [11].

3. Mathematical model

The transient model simulation used in this work contemplates a 3D gas–solid continuous flow; a 12 lump chemical catalytic cracking reactions model; inclusion of heat transfer. It is assumed that the feedstock is vaporized completely. The fluid dynamic equations were taken from the ANSYS/CFX-Solver-Theory-Guide [13], whilst the catalytic cracking kinetic models were taken from Wu et al. [10] and Chang et al. [14].

In Table 1, the governing equations for the Eulerian–Eulerian description are taken from Anderson and Jackson [15]; in the continuity equation both phases are treated as two interpenetrating continua and mass transfer between phases is not considered. In the momentum equations, the drag equation was used to model the interphase momentum transfer that is a combination of Wen and Yu correlation [16] and the Ergun equation [17]. For particle phase, when Reynolds number is large enough ($\text{Re} > 1000$) for inertial behaviors to govern viscous effects the drag constant is independent of the Reynolds number as expressed in the Eq. (19) while a low Reynolds number ($\text{Re} < 1000$), both inertial and viscous effects are significant and the drag coefficient is calculated experimentally as expressed in the Eq. (20). The modulus of elasticity (G) given by Gidaspow. [17] in Eq. (21) was used to predict the solids pressure; the packing limit is assumed to be about 0.65 for mono dispersed spheres. In the turbulence equations, the two-equation models are commonly used [4,18,19], as they offer a good compromise between numerical effort and computational precision and are much more sophisticated than the zero equation models. The k -epsilon two-equation model was used to predict the gas-phase effective turbulence viscosity as shown by Eqs. (22) and (23). Due to the energy involved in the vaporization of the liquid reactants and the endothermic reactions, the energy transfer

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