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Numerical study on gas dispersion characteristics of a coaxial mixer with viscous fluids

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A B S T R A C T

CFD (Computational fluid dynamics) method was used to study the gas dispersion characteristics of a coaxial mixer consisting of an outer anchor and an inner Rushton turbine with viscous fluids. The flow patterns of both phases and the distributions of shear strain rate, bubble diameter and local gas volume fraction were obtained. Using the total gas holdup and local gas volume fraction distribution as criteria, gas dispersion performances of the coaxial mixer operating in counter-rotation, co-rotation and single-inner-impeller rotation modes were compared. It was found that operating the mixer in counter-rotation mode exhibited better gas dispersion performance than co-rotation and single-inner-impeller rotation modes. Influences of inner impeller speed and fluid viscosity on local gas volume fraction distributions were evaluated. Results showed that increasing inner impeller speed is favorable to gas dispersion, whereas increasing fluid viscosity is not. Finally, simulated and experimental results of the vertical gas volume fraction profile in the region near the wall were compared, which showed a good agreement.

1. Introduction

Gas dispersion operations are widely encountered in food, polymerization, fermentation and sewage treatment industries. Fluids in these industrial processes are usually of high viscosity or pseudo-plastic, which imposes high requirements on the mixers. Conventional mixers are only applicable to special conditions. By contrast, coaxial mixers have the virtues of flexibility and adjustability. They can apply to a wide range of conditions, include those conditions mentioned above [1].

Previous researches concerning coaxial mixers focused mainly on their performance in single phase systems. Foucault et al. [2,3] assessed the power consumption characteristics of several coaxial mixers consisting of an outer wall-scraping impeller and different types of inner dispersing impellers operating in counter- and co-rotating modes with Newtonian and non-Newtonian fluids. They found that the speed of the outer impeller has little effect on the power draw of the dispersing impeller, whereas the speed of dispersing impeller was shown to affect the anchor power consumption significantly; the power consumption of the outer impeller would increase in counter-rotating mode and decrease in co-rotating mode. They also proposed a new correlation based on the impeller geometry for the generalized Reynolds number and the power consumption. Christian et al. [4–6] experimentally and numerically studied the power consumption, mixing time, tracer evolution, and flow characterization of coaxial mixers with different combinations of inner and outer impellers. Both co- and counter-rotation modes were investigated and results showed that operating the agitator in co-rotation mode exhibits shorter mixing time than counter-rotation mode at the same power per unit volume. Liu et al. [7,8] investigated the power consumption and heat transfer performance of a coaxial mixer operating in an agitating vessel full of malt syrup with high viscosity. By defining a new characteristic diameter and a new characteristic speed, a novel correlation for the calculation of the total power consumption of the coaxial mixer was introduced. They also found that the effect of the co-rotation mode and counter-rotation mode on heat transfer is of little difference. Pakzad et al. evaluated the effects of operating conditions (impeller speed, rotation mode, speed ratio and fluid concentration) on the performance of coaxial mixers in the mixing of yield-pseudoplastic fluids [9,10], and developed new correlations for the generalized Reynolds and power numbers of different kind of coaxial mixer configurations operating in yield-pseudoplastic fluids [11,12].

For the gas dispersion characteristics of coaxial mixers, very few investigations have been reported. Espinosa-Solares et al. [13,14] studied the gas dispersion performances of coaxial mixers consisting of a helical ribbon and either a Smith or a Rushton turbine. By observing the bubble distribution in the stirred tank, they found that coaxial mixers perform much better than individual

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Nomenclature

C_D	Drag coefficient
d_b	Bubble diameter, mm
Eo	Eötvös number
F_D	Drag force, N
g	Gravity acceleration, m s^{-2}
h	Height of monitoring point, mm
H	Height of liquid level, mm
k	Turbulent kinetic energy, m^2 s^{-2}
M_O	Morton number
Ni	Inner impeller speed, rpm
No	Outer impeller speed, rpm
Re	Reynolds number
$\bar{u}$	Velocity, m s^{-1}
Vg	Gas flow rate, m^3 h^{-1}
Greek letters	
ρ	Density, kg m^{-3}
μ	Viscosity, Pa·s
α	Volume fraction, %
σ	Surface tension coefficient, N m
ε	Turbulent kinetic energy dissipation rate, m^2 s^{-3}
subscripts	
g	gas
l	liquid

impellers in terms of gas dispersion, and coaxial mixers can also achieve good gas dispersion performance in both Newtonian and non-Newtonian fluids. To further understand the gas dispersion characteristics of coaxial mixers with viscous fluids, more investigations are necessary.

Experimental methods are usually adopted to study the gas dispersion characteristics of mixers. However, experimental studies can hardly obtain such information as the shear strain rate distribution and flow pattern in the stirred vessel, which will greatly limit the study of the gas dispersion characteristics of coaxial mixers. By contrast, the CFD method does not have such limitations, thus can make up for these shortages of the experimental method.

This work is an extension of our previous experimental research [15]. Computational fluid dynamics software — CFX 14.5 was used to study the gas dispersion characteristics of a coaxial mixer consisting of an outer anchor and an inner Rushton turbine with viscous fluids. Besides, to verify the reliability of simulation results, gas volume fractions at six points near the wall were experimentally measured and results from numerical simulation were compared with the experimental data.

2. Configuration of the mixing system

The mixing system consisting of a vessel, an outer impeller, an inner impeller, a ring sparger and coaxial shafts of inner and outer impellers is shown in Fig. 1. The vessel with a diameter of 380 mm is equipped with a standard elliptical head. The outer impeller is actually an anchor because its crossbeam does not immersed in liquid. Its outer diameter is 280 mm, and the width of its blades is 38 mm. The diameters of the Rushton turbine and its center disc are 133 mm and 90 mm, respectively. Its blade width is 27 mm. The ring sparger is located at the height of 112 mm from the bottom. It has a diameter of 120 mm and a cross section diameter of 18 mm, and 24 holes with 2 mm diameter are equispaced around its circumference.

3. Mathematical model

3.1. Equations for two-phase flow

The gas-liquid flow in the stirred vessel was modeled using the Eulerian two-fluid model, where both the continuous and the dispersed phases are treated as interpenetrating continua. The Eulerian modeling framework is based on ensemble-averaged mass and momentum transport equations for each of these phases. These transport equations can be written as:

Continuity equation:

$$\frac{\partial}{\partial t}(\alpha_k\rho_k)+\nabla(\alpha_k\rho_k\overline{u_k}) \tag{1}$$

Momentum transfer equation:

$$\begin{aligned}&\frac{\partial}{\partial t}(\alpha_k\rho_k\overline{u_k})+\nabla(\alpha_k\rho_k\overline{u_k}\ \overline{u_k})\\&=\ -\alpha_k\nabla P+\nabla\left(\alpha_k\mu_{eff,k}\left(\nabla\overline{u_k}+(\nabla\overline{u_k})^T\right)\right)+\alpha_k\rho_k g+F_{lg}\end{aligned} \tag{2}$$

where k = l for liquid, k = g for gas. P is the pressure which is shared by both phases. $\mu_{eff,k}$ is the effective viscosity and *Flg* is the ensemble-averaged momentum exchange between the two phases.

In the Eulerian frame of reference gas is treated as a continuum, so the gas phase should have the similar physical parameters such as the turbulent viscosity with the liquid. These parameters play an essential role in closing the above mass and momentum transport equations.

The effective viscosity of the liquid phase is calculated as:

$$\mu_{eff,l}=\mu_l+\mu_{t,l}+\mu_{tp} \tag{3}$$

where μ_l is the molecular viscosity of the liquid phase, $\mu_{t,l}$ is the turbulent viscosity of the liquid phase and μ_{tp} is an extra term due to bubble induced turbulence.

To obtain the turbulent viscosity of the liquid phase, the standard k-ε model was used. It assumes that the turbulent viscosity is linked to the turbulence kinetic energy and dissipation via the relation:

$$\mu_{t,l}=C_\mu\rho_l\frac{k^2}{\varepsilon} \tag{4}$$

where $C_\mu=0.09$, the values of k and ε come directly from the differential transport equations for the turbulence kinetic energy and turbulence dissipation rate:

$$\frac{\partial(\alpha_l\rho_l k)}{\partial t}+\nabla[\alpha_l(\rho_l\overline{u_l}k)]=\nabla\left[\alpha_l\left(\mu_l+\frac{\mu_{t,l}}{\sigma_k}\right)\nabla k\right]+\alpha_l(P_k-\rho_l\varepsilon) \tag{5}$$

$$\begin{aligned}\frac{\partial(\alpha_l\rho_l\varepsilon)}{\partial t}+\nabla[\alpha_l(\rho_l\overline{u_l}\varepsilon)]&=\nabla\left[\alpha_l\left(\mu_l+\frac{\mu_{t,l}}{\sigma_\varepsilon}\right)\nabla\varepsilon\right]\\&\quad+\alpha_l\frac{\varepsilon}{k}(C_{\varepsilon_1}P_k-C_{\varepsilon_2}\rho_l\varepsilon)\end{aligned} \tag{6}$$

where $C_{\varepsilon_1}, C_{\varepsilon_2}, \sigma_k$ and σ_ε are constants and their values are 1.44, 1.92, 1.0 and 1.3, respectively. P_k is the turbulence production due to viscous forces and is given by:

$$P_k=\mu_{t,l}\nabla\overline{u_l}\left(\nabla\overline{u_l}+(\nabla\overline{u_l})^T\right)-\frac{2}{3}\nabla\overline{u_l}\left(3\mu_{t,l}\nabla\overline{u_l}+\rho_l k\right) \tag{7}$$

In addition, to calculate the extra term, μ_{tp}, the model proposed by Sato and Sekoguchi [16] was adopted, and therefore:

$$\mu_{\mathrm{tp}}=C_{\mu\mathrm{P}}\mu_l\alpha_g d_b|\overline{u_g}-\overline{u_l}| \tag{8}$$

with a model constant $C_{\mu\mathrm{P}}=0.6$.

The effective gas viscosity is formulated as:

$$\mu_{eff,g}=\mu_g+\mu_{t,g} \tag{9}$$