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CFD INVESTIGATION OF SLUGGING BEHAVIOR IN A GAS-SOLIDS FLUIDIZED BED

Article Highlights

- A simple two-fluid model is applied to simulate the slugging behavior
- Effect of viscous force on slugging dynamics is investigated
- Predicted results are in good agreement with the experimental findings

Abstract

The hydrodynamics of a slugging fluidized bed is investigated numerically by using a two-fluid model suggested by Brandani and Zhang [22]. Numerical simulations are carried out in the platform of a commercial software package, CFX 4.4, by adding user-defined subroutines. The bed expansion ratio, the pressure drop fluctuation and its power spectrum density at the equipment scale and the solids volume fraction distributions at the grid scale are predicted to explore the slugging fluidization characteristics for Geldart group D particles. By comparing with the experimental data in the literature, the numerical model is found to have better predictive ability when the viscous force term is considered in the gas phase momentum equation.

Keywords: gas-solid fluidized bed, slugging, Geldart group D particles, CFD simulation.

Gas-solid fluidized beds have been extensively used in petroleum refining and petrochemical processes because of their excellent heat and mass transfer properties, and effective utilization of active catalytic surface. In parallel with industrial developments, many academic efforts have been focused on the fluidization for providing a theoretical framework to strengthen this subject [1]. Grace *et al.* [2] proposed a flow regime map, including homogeneous, bubbling, slugging, turbulent and fast fluidization. Up to now, the majority of studies on gas-solids fluidized beds were centered on the bubbling or fast fluidization behaviors, which are common in industrial process.

The slugging is characterized as the size of bubble becomes almost the same as the diameter of fluidized bed [3]. It can be explained for bubbling beds by considering that the bubble growth reaches the

bed diameter [4]. The slugging behavior occurs easily when the large particles, especially for Geldart group D particles, are employed, which should result in more unstable fluidization patterns [5]. For different types of fluidized beds, sometimes the slugging is encountered in the bed of high aspect ratio, often with a small diameter compared to the particle size [6]. Generally, the slugging is categorized into three types: round-nosed slug, wall slug, and square-nosed slug. Wu *et al.* [7] found that the fluidized beds with Geldart group D particles could operate in the round-nosed slug, whilst the wall slug or square-nosed slug formed with the increase in superficial gas velocity or static bed height. As it forms a complex and heterogeneous structure traversed by voids or bubbles, the slugging behavior is usually considered to be unfavorable to the performance of fluidized bed reactor [8]. Therefore, it is necessary to understand the slugging characteristics for developing and scaling up the fluidized-bed reactors in the field of process industry.

Computational fluid dynamics (CFD) modeling has been extended successfully to investigate the hydrodynamics of gas-solids systems. In general, CFD models are divided into two categories: Eulerian-Lagrangian and Eulerian-Eulerian approaches. The

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former considers the solids phase at the particle level, which is still too complicated to be applied in an engineering installation. While the latter considers both fluid phase and solids phase as two inter-penetrating continua, and the two phases are described as separate conservation equations with appropriate interaction terms, which can handle the large-scale fluidized bed since the point variables are averaged over local region. The Eulerian-Eulerian model, also called two-fluid model (TFM), has been suggested as a feasible approach for performing parametric investigations and scale-up studies in the fluidized beds [9]. A set of physical or empirical models are employed to close the conservation equations in the TFM, especially for the solids phase stress term in the momentum equations [10]. However, there is still no consensus on the particle-phase viscosity in the solids momentum equation for CFD simulations based on TFM model [11,12]. The solids-phase viscosity is generally defined by three approaches: *i)* as a constant of 1.0 based on experimental data of Grace [13], *ii)* as a constant value of zero and *iii)* as a function of the granular temperature by kinetic theory of granular flow (KTGF) [14]. In recent twenty years, a few investigations were carried out for studying the gas-solids slugging fluidization based on the TFM together with KTGF [15–20], such as Pain *et al.* [15] investigated the formation, elongation, coalescence and eruption of bubbles in the slugging fluidized bed. Zhang and Yu [16] obtained the effect of wall conditions on slugging formation. Lettieri *et al.* [1,17] simulated the behavior of fluidized beds under different flow regimes, spanning from bubbling to slugging and turbulent fluidization.

Based on the inviscid two-fluid model of Gidaspow [21], Brandani and Zhang [22] proposed a hydrodynamic model by introducing additional terms into both gas- and solid-phases momentum balance equations. Series of simulations were implemented to explore the fluidization characteristics, such as homogeneous fluidization of Geldart group A particles, bubbling and collapsing behavior for Geldart group B particles, and the pressure fluctuation characteristics in the gas-solid fluidized bed [23–25]. The gas viscosity term was ignored in the gas phase momentum equation when an additional term was taken into account and the relatively low operational gas velocity was used in the homogeneous or bubbling fluidization regime. The effect of the gas viscosity term in the gas phase momentum equation is explored to predict the gas-solid slugging fluidization behaviors, such as solids volume fraction, bed expansion, and pressure drop fluctuation.

CFD MODEL AND NUMERICAL PROCEDURE

Mathematical model

The continuity and momentum balance equations, describing gas and particle flows in the two-dimensional cold model of fluidized bed, are given below:

Continuity equations:

Gas phase:

$$\frac{\partial \varepsilon_g}{\partial t} + \nabla \cdot (\varepsilon_g \vec{u}_g) = 0 \quad (1)$$

Particle phase:

$$\frac{\partial \varepsilon_p}{\partial t} + \nabla \cdot (\varepsilon_p \vec{u}_p) = 0 \quad (2)$$

Momentum equations:

Gas phase:

$$\begin{aligned} \frac{\partial (\varepsilon_g \rho_g \vec{u}_g)}{\partial t} + \nabla \cdot (\varepsilon_g \rho_g \vec{u}_g \vec{u}_g) = \\ = -\beta (\vec{u}_g - \vec{u}_p) - \varepsilon_g \nabla p + \varepsilon_g \rho_g \vec{g} - \vec{F}_{ad,g} \end{aligned} \quad (3)$$

Particle phase:

$$\begin{aligned} \frac{\partial (\varepsilon_p \rho_p \vec{u}_p)}{\partial t} + \nabla \cdot (\varepsilon_p \rho_p \vec{u}_p \vec{u}_p) = \\ = -\beta (\vec{u}_p - \vec{u}_g) - \varepsilon_p \nabla p + \varepsilon_p \rho_p \vec{g} - \vec{F}_{ad,p} \end{aligned} \quad (4)$$

where ε is the volume fraction ($\varepsilon_g + \varepsilon_p = 1$), t the time, ρ the density, \vec{u} the velocity vector, p the pressure, β the inter-phase drag coefficient, \vec{g} the acceleration due to gravity, and \vec{F}_{ad} the additional force. The subscripts g and p indicate gas and particle phases, respectively.

The inter-phase momentum transfer plays an important role in modeling the particle-fluid interactions. And the inter-phase drag term is a function of the particle drag coefficient, C_D , which can be derived as follows:

$$\beta = C_D \frac{3\varepsilon_p \rho_g |\vec{u}_p - \vec{u}_g|}{4d_p} \varepsilon_g^{-1.8} \quad (5)$$

C_D can be obtained from the empirical Dalla Valle relationship [26]:

$$C_D = \left(0.63 + \frac{4.8}{\sqrt{Re}} \right)^2 \quad (6)$$

here:

$$Re = \frac{\varepsilon_g \rho_g |\vec{u}_p - \vec{u}_g| d_p}{\mu_g} \quad (7)$$

where d_p is the particle diameter, μ is the dynamic viscosity, and Re is the particle Reynolds number.

Based on a linear stability analysis of Wallis [27] and Gibilaro [28], the additional forces, \vec{F}_{ad} , are introduced into the inviscid two-fluid model of Gidaspow [21] by Brandani and Zhang [22]. By compared experimental with predicted minimum bubbling points [29-31], the characteristic length for describing the discrete nature of the particles in this additional term is of the order of particle diameter. The above method can be used to predict the homogeneous, bubbling, collapsing or jetting behaviors for Geldart group A or B particles in the gas-solids fluidized bed [23-25] and to explore the liquid-solid fluidization [32,33]. Due to the superficial gas velocity for the slugging phenomenon is much higher than that for the homogeneous or bubbling state in the gas-solids fluidized bed, the effect of gas viscosity term in the gas momentum equation is introduced into the additional force term, $\vec{F}_{ad,g}$ in the TFM suggested by Brandani and Zhang [22]. Accordingly, the additional forces are expressed as:

$$\vec{F}_{ad,g} = d_p \left[(1 - 2\varepsilon_g) \rho_p + 2\varepsilon_g \rho_g \right] \vec{g} \nabla \varepsilon_g \vec{i} + \nabla(\vec{\tau}_g) \quad (8)$$

$$\vec{F}_{ad,p} = d_p \left[2\varepsilon_p \rho_p + (1 - 2\varepsilon_p) \rho_g \right] \vec{g} \nabla \varepsilon_g \vec{i} \quad (9)$$

where \vec{i} is unit vector, which is the same direction as acceleration \vec{g} , $\vec{\tau}_g$ is the viscosity tension.

Numerical procedure

In order to save computational resources, a detailed CFD investigation is implemented in the 1.0 m (height) × 0.1 m (width) 2-D fluidized bed. The structure grid of 160 (height) × 16 (width) is employed based on the grid independency analysis, and time step is selected as 2×10^{-4} s. In order to validate the reliability of the present numerical model, the simulated results are compared with the experimental data reported by Yang *et al.* [18], which were obtained in the fluidized bed with an inner diameter of 0.1 m and a height of 1.0 m. The physical properties of gas-solids phases are listed in Table 1.

Table 1. Physical properties of gas-solids system

Parameter	Value
Gas density, ρ_g , kg/m ³	1.225
Particle density, ρ_p , kg/m ³	1294
Particle diameter, d_p , μ m	2200
Minimum fluidization velocity, u_{mf} , m/s	0.6
Minimum voidage, ε_{mf}	0.4
Initial bed height, H_0 , m	0.3

The boundary conditions in the 2-D fluidized bed are: *i*) in the bottom of bed, the gas inlet velocity is specified and the solids velocity is set to zero in the vertical direction. The gas and solids velocities are zero in the horizontal direction; *ii*) in the top of bed, the constant static pressure boundary is adopted, *i.e.*, ambient atmosphere; *iii*) left and right walls are treated as the no-slip velocity boundary conditions for both gas and solid phases; *iv*) front and back walls are regards as symmetry planes since the simulations are carried out in the 2-D bed.

Numerical calculation is performed on the platform of a commercial CFD code, CFX 4.4. The continuity equations (Eqs. (1) and (2)) are solved directly by the software. In momentum equations (Eqs. (3) and (4)), the pressure drop and gravity force terms are defined in the command file, while the inter-phase drag and additional force terms in Eqs. (5), (8) and (9) are programmed by user-defined Fortran subroutines. The numerical solutions of the modeling equations are obtained by a finite volume method based on a collocated grid approach by used the Rhie-Chow algorithm [34] to prevent checkerboard oscillations. SIMPLEC algorithm (semi-implicit method for pressure linked equations-consistent algorithm) is used to deal with the pressure-velocity coupling. Under-relaxation factors are 0.65 for gas (or solid) volume fraction and velocity vector, and 1.0 for pressure. Different differencing methods are used to treat the advection terms: central differencing scheme for gas or solids volume fraction, upwind differencing scheme for the shared pressure field, and hybrid differencing scheme for all velocity components.

Initially, the bed keeps in the minimum fluidized state, and the initial bed height is set to H_0 . The volume fraction of solids, ε_p , should be zero in the dilute phase region above bed surface. However, ε_p is set to 10^{-10} during the practical computational procedure, so as to provide more realistic results for the particle velocity and ensure good convergence. In the dense phase region, the particles are uniformly suspended at the minimum fluidization state. The initial pressure profile is calculated from the hydrostatic bed height based on the reference pressure:

$$P = P_0 + \left[\varepsilon_{mf} \rho_g + (1 - \varepsilon_{mf}) \rho_p \right] (H_0 - H) \vec{g} \quad (10)$$

RESULTS AND DISCUSSION

Determination of the minimum slugging velocity

The formation and development of slugging are depend upon the geometrical structure of fluidized bed, physical properties of gas-solid system, and

operating condition. When the geometrical structure and physical properties are determined, the superficial gas velocity plays an important role for slugging behaviors. Figure 1 shows the instantaneous distributions of solids volume fraction at different velocities. When the superficial gas velocity is less than 0.4 m/s, the bed height keeps constant either introducing gas viscosity item (GVI) in Eq. (8) or not, since the bed maintains in fixed-bed state. The bed enters fluidization state when the gas velocity is increased to 0.6 m/s. A bubble appears in the lower part of the bed when the GVI is ignored, but a big bubble (slug) can be observed in the upper part of the bed when the GVI is added. As the gas velocity is further increased to 0.8 m/s, distinct slugs are observed within the fluidized beds either considering GVI or ignoring GVI.

Based on experimental data, Stewart and Davidson [3] proposed an experiential correlation for predicting the minimum slugging velocity as below:

$$u_{msl} = u_{mf} + 0.07(gD)^{0.5} \quad (11)$$

where u_{msl} is the minimum slugging velocity, u_{mf} the minimum fluidization velocity, g the gravitational acceleration, and D the bed diameter.

For the gas-solid system and fluidized bed used in this study, the minimum slugging gas velocity is calculated as 0.67 m/s, which is close to the numerical value of 0.6 m/s with GVI or 0.8 m/s without GVI.

Solids volume fraction

Figure 2 indicates the distribution of solids volume fraction obtained at the gas velocity of 1.2 m/s (twice of minimum fluidization velocity). The rounded slugs are obtained when GVI is not taken into account (Figure 2a), however obvious wall slugs are formed when GVI is considered (Figure 2b). The wave of bubbles appears in the lower part of bed and travels through the bed, which leads to a noticeable

bed expansion. From the snapshots of solids volume fraction with small time interval, it is found that the bubbles move upward and then merge into a dilute flow core in the upper part of the bed, while particles at the top of the slug fall back to the dense region in the lower part of the bed. The results from Yang *et al.* [18] showed an increase in bubble size with increasing the superficial gas velocity, and the distinct wall slug occur as the gas velocity higher than 0.8 m/s. CFD simulation in this study indicates that a stable and regular slugging is formed within the fluidized bed, which is in good agreement with the findings of Yang *et al.* [18].

Bed expansion

Bed expansion process is mainly affected by the movement of bubbles for Geldart group D particles, and it has an ability to describe the characteristics of bubbles in the fluidized bed. Accordingly, the time-averaged bed expansion ratio is one of the important parameters for scaling up the fluidized bed. Figure 3 compares experimental and computational bed expansion ratios. Initially, the bed expansion ratio is set as 1 in the fixed bed state; the computational results show a reasonably match with the experimental values owing to the superficial gas velocity is less than the minimum fluidization velocity. The bed expansion ratio increases with increasing the superficial gas velocity, when the superficial gas velocity is greater than the minimum fluidization velocity. The simulated bed expansion ratios with GVI or without GVI are close to the experimental results [18] within a relative error of 12%.

Pressure drop fluctuation and power spectrum density

Pressure drop fluctuation derives from bubble movement in the gas-solids fluidized bed. Figure 4 exhibits the effect of GVI on the pressure drop at $u_g = 1.2$ m/s. Bubble (or slug) phase and emulsion phase are distinguished, *i.e.*, the peaks are considered as

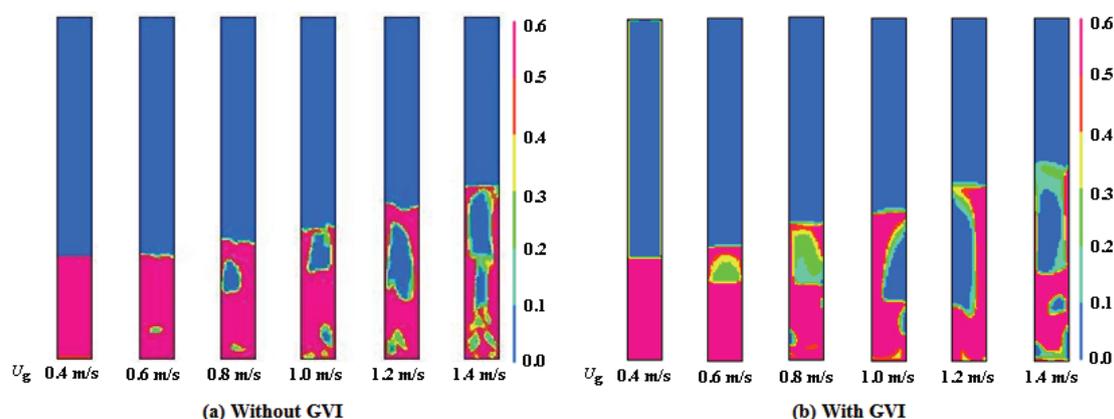


Figure 1. Instantaneous distribution of solids volume fraction at different gas velocities.

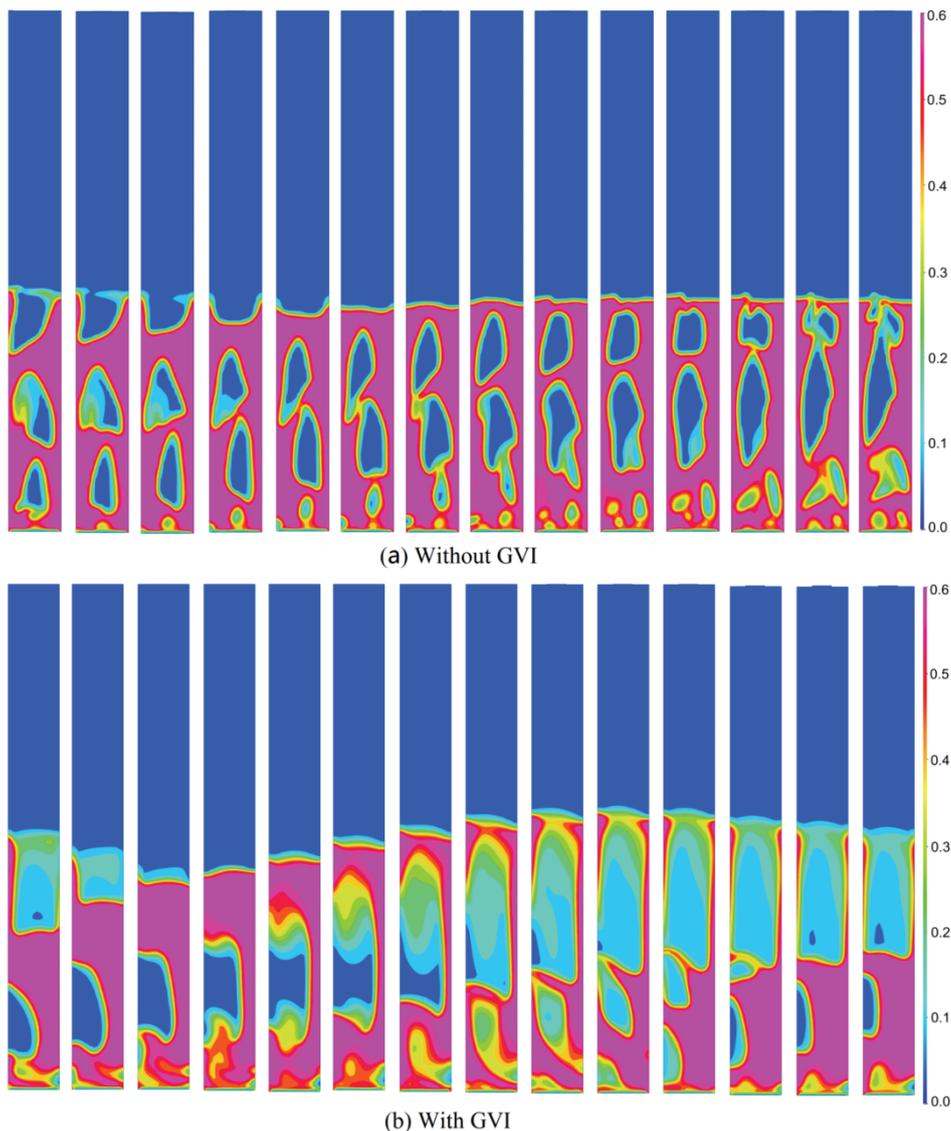


Figure 2. Snapshots of solids volume fraction.

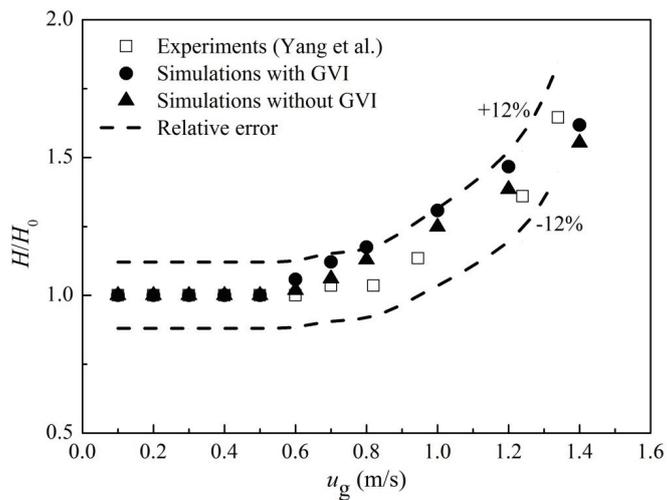


Figure 3. Comparison of simulated and experimental time-averaged bed expansion ratios.

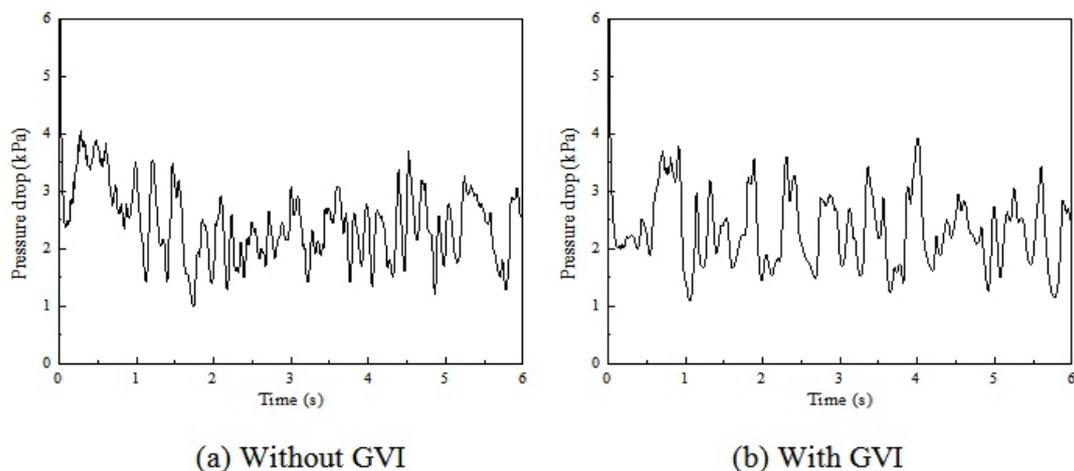


Figure 4. Effect of gas viscosity term on the bed pressure drop fluctuation, $u_g = 1.20$ m/s.

the dense phase, whilst the troughs signify the bubbles or slug region. Furthermore, the time-averaged pressure drops predicted are about 2348.5 Pa by ignored GVI and 2326.2 Pa by considered GVI, which are close to the experimental data of Yang *et al.* [18].

Figure 5 presents the power spectral density obtained with a data acquisition frequency of 100 Hz at the superficial gas velocity of 1.2 m/s. The gas velocity has a great influence on the power spectrum of the bed pressure drop fluctuations. The energy distribution is more uniform when GVI is not taken into account, which is similar to that in the bubbling bed (Figure 5a). However, the maximum power spectrum is obtained in the low-frequency region with higher energy when GVI is added in Eq. (8) (Figure 5b). An obvious peak frequency of 1.87 Hz is found close to the experimental result of 1.36 Hz [18], while the energy is relatively small in other high-frequency parts. Such a phenomenon represents the slug cycle of formation, development and breakup, which is corresponding to the snapshots of solids volume fraction

in Figure 2b. Qualitative and quantitative numerical results with considering GVI show good agreement with the experimental data of Yang *et al.* [18].

CONCLUSIONS

The slugging behavior is investigated in a gas-solid fluidized bed by using a hydrodynamic model proposed by Brandani and Zhang [22]. The main conclusions are: *i*) the gas viscosity item in the momentum equation has an effect on the slugging formation in the gas-solid fluidized bed; *ii*) simulated minimum slugging gas velocity of 0.6 m/s (or 0.8 m/s) with (or without) gas viscosity term is close to the prediction of Stewart and Davidson [3]; *iii*) the computational bed expansion ratios are in agreement with the experimental results of Yang *et al.* [18]; *iv*) the analysis of the power spectral density shows that a higher energy maximum is found in the low-frequency region with the gas viscosity term, which is in good agreement with the experimental findings.

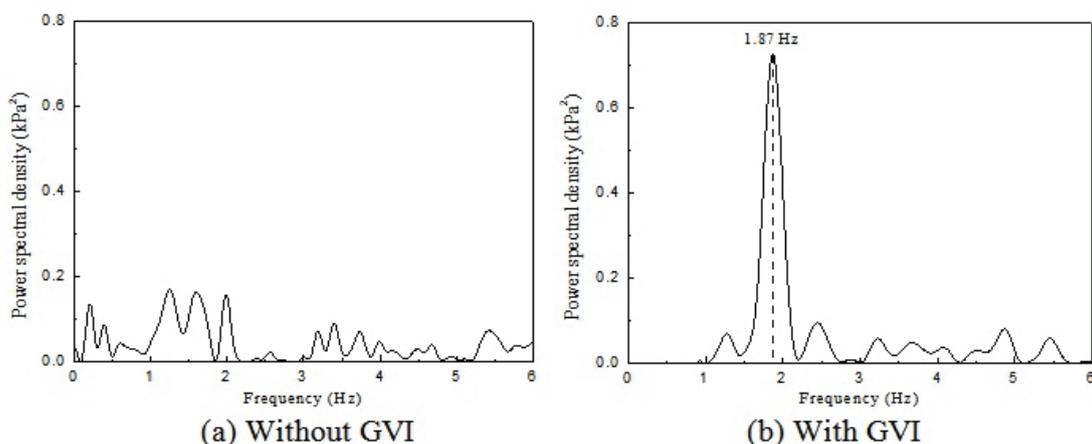


Figure 5. Power spectrum density of the bed pressure drop fluctuation signals, $u_g = 1.20$ m/s.

Nomenclature

C_D	particle drag force coefficient
d_p	particle diameter, m
\vec{F}_{ad}	additional force, N/m ³
\vec{g}	acceleration due to gravity, m/s ²
H	bed height, m
H_0	static bed height, m
P	bed pressure, N/m ²
P_0	atmospheric pressure, N/m ²
ρ	pressure, N/m ²
Re	Reynolds number
t	time, s
\vec{u}	velocity vector, m/s
u_{mf}	minimum fluidization velocity, m/s

Greek letters

β	interphase drag coefficient
ε	volume fraction
ε_{mf}	minimum voidage
μ	viscosity, Pa·s
ρ	density, kg/m ³

Subscripts

g	gas phase
p	particle phase
x	vector component in x direction
z	vector component in z direction

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REFERENCES

- [1] P. Lettieri, G. Saccone, L. Cammarata, *Chem. Eng. Res. Des.* **82** (2004) 939-944
- [2] J. Grace, *Powder Technol.* **113** (2000) 242-248
- [3] P. Stewart, J. Davidson, *Powder Technol.* **1** (1967) 61-80
- [4] D. Gidaspow, P. Chaiwang, *Chem. Eng. Sci.* **97** (2013) 152-161
- [5] Y. Jin, J. Zhu, Z. Wang, Z. Yu, *Fluidization Engineering Principles* (in Chinese), Tsinghua University Press, Beijing, 2001, p. 85
- [6] P. Lettieri, R. di Felice, R. Pacciani, O. Owoyemi, *Powder Technol.* **167** (2006) 94-103
- [7] G. Wu, J. Ouyang, B. Yang, Q. Li, F. Wang, *Particuology* **10** (2012) 72-78
- [8] X. Ku, T. Li, T. Lovas, *Chem. Eng. Sci.* **95** (2013) 94-106
- [9] B. van Wachem, J. Schouten, C. van den Bleek, R. Krishna, J. Sinclair, *AIChE J.* **47** (2001) 1035-1051
- [10] S. Hosseini, W. Zhong, M. Esfahany, L. Pourjafar, S. Azizi, *J. Fluids Eng.* **132** (2010) 041301
- [11] J. Li, J. Ouyang, S. Gao, W. Ge, N. Yang, W. Song, *Multi-scale simulation of particle-fluid complex-fluid complex systems* (in Chinese), Science Press, Beijing, 2005
- [12] D. Patil, M. van Sint Annaland, J.A.M. Kuipers, *Chem. Eng. Sci.* **60** (2005), 57-72
- [13] J.R. Grace, *Can. J. Chem. Eng.* **48** (1970) 30-33
- [14] K. Zhang, P. Pei, S. Brandani, H. Chen, Y. Yang, *Chem. Eng. Sci.* **68** (2012) 108-119
- [15] C. Pain, S. Mansoorzadeh, C. de Oliveira, *Int. J. Multiphas. Flow* **27** (2001) 527-551
- [16] S. Zhang, A. Yu, *Powder Technol.* **123** (2002) 147-165
- [17] P. Lettieri, L. Cammarata, G. Micale, J. Yates, *Int. J. Chem. React. Eng.* **1** (2003) 1-19
- [18] F. Yang, J. Wang, X. Gu, *Chin. J. Pro. Eng.* **5** (2005) 597-600
- [19] N. Ratkovich, C. Chan, P. Berube, I. Nopens, *Chem. Eng. Sci.* **64** (2009) 3576-3584
- [20] X. Wang, L. Yu, J. Wang, *Int. J. Chem. React. Eng.* **10** (2012) A77
- [21] D. Gidaspow, *Multiphase Flow and Fluidization*, Academic Press, Boston, MA, 1994
- [22] S. Brandani, K. Zhang, *Powder Technol.* **163** (2006) 80-87
- [23] K. Zhang, S. Brandani, J. Bi, J. Jiang, *Prog. Nat. Sci.* **18** (2008) 729-733
- [24] P. Pei, K. Zhang, E. Lu, D. Wen, *Petrol. Sci.* **6** (2009) 69-75
- [25] Q. Wang, K. Zhang, H. Gu, *Petrol. Sci.* **8** (2011) 211-218
- [26] J. Dalla Valle, *Micromeritics*, Pitman, London, 1948
- [27] G. Wallis, *One-Dimensional Two-Phase Flow*, McGraw-Hill, New York, 1969
- [28] L. Gibilaro, *Fluidization Dynamics*, Butterworth Heine-mann, London, 2001
- [29] K. Rietema, H.W. Piepers, *Chem. Eng. Sci.* **45** (1990) 1627-1639
- [30] A. Marzocchella, P. Salatino, *AIChE J.* **46** (2000) 901-910
- [31] K. Godard, J.F. Richardson, *I. Chem. E. Symp. Ser.* **30** (1968) 126-135
- [32] K. Zhang, G. Wu, S. Brandani, H. Chen, Y. Yang, *Powder Technol.* **227** (2012) 104-110
- [33] K. Zhang, Y. Guan, X. Yao, Y. Li, X. Fan, S. Brandani, *Powder Technol.* **235** (2013) 180-191
- [34] C. Rhie, W. Chow, *AIAA J.* **21** (1983) 1525-1532.

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NAUČNI RAD

CFD ISTRAŽIVANJE PONAŠANJA ČEPOVA U FLUIDIZOVANOM SLOJU GASNO-ČVRSTO

Hidrodinamika fluidizovanog sloja sa čepovima je istraživan korišćenjem modela sa dva fluida koji su predložili Brandani i Zhang. Numeričke simulacije su izvedene na platformi komercijalnog softverskog paketa CFX 4.4, dodavanjem korisničkih potprograma. Model proračunava odnos ekspanzije sloja, fluktuaciju pada pritiska i gustinu spektra snage na nivou opreme i raspodela sadržaja čvrste faze na nivou mreže u cilju određivanja karakteristika fluidizovanog sloja sa čepovima u slučaju za čestica Geldart group D. Upoređivanjem sa eksperimentalnim podacima iz literature, utvrđeno je da numerički model ima bolju prediktivnu sposobnost kada se član viskozne sile uključi u jednačini količine kretanja za gasnu fazu.

Ključne reči: fluidizovani sloj gas-čvrsta materija, formiranje čepova, Geldartove čestice, CFD simulacija.