STUDY OF THE SCALE-UP RELATIONS FOR SPOUTED BEDS USING CFD

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Abstract. *Spouted beds have been used in several engineering applications including drying, coating, granulation and gas-solid chemical reactions. However, this equipment presents some operational and design problems during scale-up, which limit the capacity of the equipment and its industrial application. Some papers in literature have already presented several scale-up relations for the spouted bed. However, this subject is still controversial and demanding further research. From the availability of CFD (Computational Fluid Dynamics) techniques, rigorous simulation of the transport phenomena present in the spouted bed can be used to evaluate scale-up procedures without the necessity to set-up a spouted bed. Therefore, this work analyzes if these different scale-up relations present in the literature for spouted beds can be used extensively for scaling up. The fluid dynamics and solids behaviors in the spouted bed were obtained through simulations and these simulations are compared with experimental data for verification. It was numerically verified that spouted bed scale-up relations presented in the literature give good results for equipments of low capacity when all the relations are satisfied. It was also verified the behavior of a scale-up procedure for an industrial case, where the solid and fluid characteristics are considered constant. In this case, the spouted bed scaled-up did not present the same fluid dynamics and solids behaviors. Instead, an unstable behaviour was observed with larger scales. The commercial package FLUENT 6.2.16 was used for the simulations.*

Keywords*: spouted bed, simulation, CFD, scale-up*

1. INTRODUCTION

Spouted beds were originally invented in Canada by Mathur and Gishler in 1955, arising as an alternative to fluidized beds for handling coarse particles. The circulation of solids inside the system makes the spouted bed an equipment that promotes high rates of mass and heat transfer. Due to this characteristic, spouted beds are now widely applied in several physical and chemical operations, such as drying (Benali and Amazous, 2006; Jenkins *et al.*, 2002), coating (Jono *et al.*, 2000; Ichikawa *et al.*, 2003), carbonization (Salam and Battacharya, 2006a; Rasul, 2001), and pyrolysis (Aguado *et al.*, 2003).

Many equations based on small-scale vessels with column diameter (*Dc*) up to 0.3m are available in the literature for predicting the fluid dynamics properties of spouted beds. Although the fluid mechanics is crucial to the proper design of large beds, there is considerable uncertainty with respect to the scale-up for this equipment. For larger spouted beds, most of existing equations do not work well. The study of scale-up for spouted beds has a great importance, because it makes possible the increase of production rate and, therefore, the number of its application in industry can also be extended.

The principle of similarity is often used in obtaining experimental data to represent large-scale complex flow phenomena. The basic concept is that if the flow fields are geometrically similar and operated with identical values of all important independent non-dimensional parameters, then the dependent non-dimensional variables must also be identical at corresponding locations (Bisio and Kabel, 1985).

Glicksman (1984) developed systematically a set of scaling laws for the proper design of fluidized beds. The set of general independent dimensionless parameters indicated by this author can be used to characterize flow regimes and fluid dynamic behavior for fluidized beds. He *et al.* (1997) examined whether the scaling relations proposed by Glicksman (1984) can be adapted to spouted bed scale-up. He *et al.* (1997), after analyzing the force balance on particles in the annulus region of spouted beds, added two additional non-dimensional parameters in the Glicksman (1984) relations: the internal friction angle and the loose packed voidage. He *et al.* (1997) also presents a detailed experimental study applying this set of relations for different scales of spouted beds, proving the validity of them for spouted beds with small sizes.

The main interest from an industrial point of view is to make the scale-up of spouted beds by keeping the same particle and fluid during the scale-up procedure. The scale-up relations of He *et al.* (1997) indicates that this procedure can not keep the full set of scaling relations unchanged. The agreement of all relations proposed by He *et al.* (1997) implies that the fluid or the particles physical characteristics, or both, must be different for different scales. The relation D_c/d_s also imposes that, for different column diameter values, different particle diameters (d_s) values are also necessary.

In recent years, numerical simulation has been widely used for studying gas-particle systems. The main advantage of a numerical simulation is that wide range of flow properties may be obtained simultaneously (Takeuchi *et al.*, 2004).

Computational Fluid Dynamics (CFD) is one of the most productive numerical methods, which can help to shorten product and process development cycles and to optimize process operation by improving energy efficiency and environmental performance (Szafran and Kmiec, 2004). In addition to several experimental about spouted beds, the use of CFD techniques in this system have brought the attention of many researchers into this field (Huilin *et al.*, 2004; Szafran and Kmiec, 2004; Duarte *et al.*, 2005; Du *et al.*, 2006).

In this work, the commercial program FLUENT 6.2.16 was used to carry out computer simulations for spouted beds, at several scales and operational conditions, using the scaling relations proposed by He *et al.* (1997). The Eulerian-Eulerian modeling approach was applied to predict the flow behavior.

2. MATHEMATICAL MODELING

2.1. The spouted bed

A stable operation of a spouted bed is characterized by a spout, fountain and annulus regions. Figure 1 shows these regions.

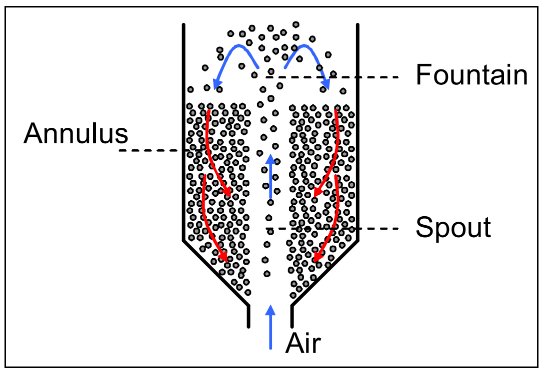


Fig. 1 – Scheme of a conventional spouted bed

The spout is a dilute solid phase, where particles are carried upward by the central jet fluid. In the annulus, the solid phase is dense and the particles and fluid move in countercurrent flow. In the fountain region, particles coming from the spout change their motion falling back into the annulus (Passos *et al.*, 1994).

In the Euler-Euler approach, the different phases are treated as interpenetrating continua, and the concept of phase volume fraction is introduced. Using an Eulerian model, a set of n momentum and continuity equations are solved for each phase.

2.2. Conservation equations of mass and momentum

The volume of the phase q (solid, fluid) is defined by:

$$V_q = \int_V \alpha_q dV \tag{1}$$

where

$$\sum_{q=1}^{n} \alpha_q = 1 \tag{2}$$

The volume fraction of each phase is calculated using the continuity equation:

$$\frac{\partial}{\partial t}\left(\alpha_q \rho_q\right) + \nabla \cdot \left(\alpha_q \rho_q \vec{v}_q\right) = 0 \tag{3}$$

The momentum conservation balance for the solid phase is:

$$\frac{\partial}{\partial t}(\alpha_s \rho_s \vec{v}_s) + \nabla \cdot (\alpha_s \rho_s \vec{v}_s \vec{v}_s) = -\alpha_s \nabla p - \nabla p_s + \nabla \cdot \overline{\overline{\tau}}_s + \alpha_s \rho_s \vec{g} + \sum_{p=1}^{n} K_{gs}(\vec{v}_g - \vec{v}_s) + (\vec{F}_s + \vec{F}_{lift,s} + \vec{F}_{vm,s}) \tag{4}$$

where:

$\vec{F}_s$ - external body force;
$\vec{F}_{lift,s}$ - lift force;
$\vec{F}_{vm,s}$ - virtual mass force;
p - pressure shared by all phases;
p_s - solid pressure;
$\overline{\overline{\tau}}_s$ - solid phase stress-strain tensor;
K_{gs} - interphase momentum exchange coefficient.

In general, the drag force is much higher than the lift force. In this case, lift force was neglected. The virtual mass effect occurs when the secondary phase *p* accelerates relative to the primary phase *q*. It is important when the secondary phase density is much smaller than the primary phase density. In this case, the virtual mass effect was also neglected.

Using the Gidaspow (1992) model, the interphase exchange coefficient for a gas-solid system can be written in the following form:

$$K_{gs} = \frac{3}{4} C_D \frac{\alpha_s \alpha_g \rho_g |\vec{v}_s - \vec{v}_g|}{d_s} \alpha_g^{-2.65}, \ \alpha_g > 0.8 \tag{5}$$

$$K_{gs} = 150 \frac{\alpha_s (1-\alpha_g)\mu_g}{\alpha_g d_s^2} \alpha_g^{-2.65} + 1.75 \frac{\rho_g \alpha_s |\vec{v}_s - \vec{v}_g|}{d_s}, \alpha_g \le 0.8 \tag{6}$$

The drag coefficient (C_D) is defined as:

$$C_D = \frac{24}{\alpha_g \mathrm{Re}_s} \left[1 + 0.15(\alpha_g \mathrm{Re}_s)0.687\right] \tag{7}$$

where the relative Reynolds number (Re_s) is calculated from

$$\mathrm{Re}_s = \frac{\rho_g d_s |\vec{v}_s - \vec{v}_g|}{\mu_g} \tag{8}$$

The solid pressure, given by Lun *et al.* (1984), is composed of a kinetic term and a second term due to the particle collisions:

$$p_s = \alpha_s \rho_s \theta_s + 2\rho_s (1 + e_{ss}) \alpha_s^2 g_{0,ss} \theta_s \tag{9}$$

In Eq. (9), e_{ss} is the coefficient of restitution for particle collisions and $g_{0,ss}$ represents the radial distribution function. In the literature, there are several formulations for the radial distribution function. In this case, $g_{0,ss}$ was expressed according to Ogawa *et al.* (1980):

$$g_{0,ss} = \left[1 - \left(\frac{\alpha_s}{\alpha_{s,\max}}\right)^{1/3}\right]^{-1} \tag{10}$$

The granular temperature for the solid phase (θ_s) is proportional to the kinetic energy of particles. It can be obtained from Ding and Gidaspow (1990):

$$\frac{3}{2}\left[\frac{\partial}{\partial t}(\rho_s\alpha_s\theta_s)+\nabla\cdot(\rho_s\alpha_s\vec{v}_s\theta_s)\right]=\left(-p_s\bar{\bar{I}}+\bar{\bar{\tau}}_s\right):\nabla\vec{v}_s+\nabla\cdot\left(k_{\theta_s}\nabla\theta_s\right)-\gamma_{\theta_s}+\phi_{gs} \tag{11}$$

where:

$\left(-p_s\bar{\bar{I}}+\bar{\bar{\tau}}_s\right):\nabla\vec{v}_s$ - generation of energy by the solid stress tensor

$k_{\theta_s}\nabla\theta_s$ - diffusion of energy

γ_{θ_s} - collisional dissipation of energy

ϕ_{gs} - energy exchange between the fluid phase and solid phase

The collisional dissipation of energy is represented by the expression derived by Lun *et al.* (1984):

$$\gamma_{\theta_s}=\frac{12\left(1-e_{ss}^2\right)g_{0,ss}}{d_s\sqrt{\pi}}\rho_s\alpha_s^2\theta_s^{3/2} \tag{12}$$

The energy exchange between the fluid phase and solid phase due to the kinetic energy of random fluctuations in particle velocity is obtained from Gidaspow *et al.* (1992):

$$\phi_{gs}=-3K_{gs}\theta_s \tag{13}$$

3. SOLUTION PROCEDURE

The CFD simulations were conducted using the commercial package FLUENT 6.2.16. A 2D axisymmetric segregated solver was chosen, and simulations of only half column were carried out to reduce the computational effort. The grid contains triangular cells in the cone base and square cells in the cylindrical region, as can be verified in Fig. 1. The mesh was composed by grid spaces of 5% larger than the particle diameter. A finite control volume technique was used to solve the set of governing equations, with the pressure-velocity coupling obtained by SIMPLE algorithm, with typical values of underrelaxation factors between 0.2 and 0.4. Solution was considered converged when the residuals of all variables were less than $1.0e^{-3}$. The steady-state was reached after 8 seconds of real time, using time intervals between $1.0e^{-4}$s and $1.0e^{-3}$s.

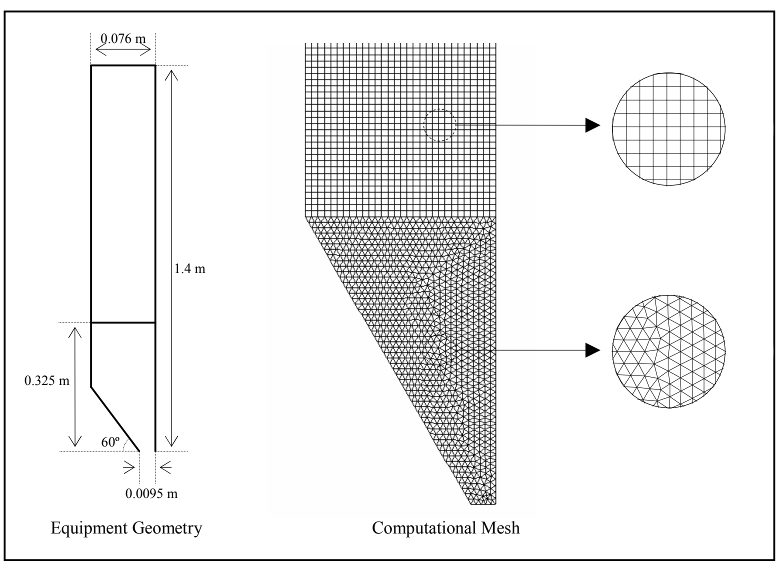


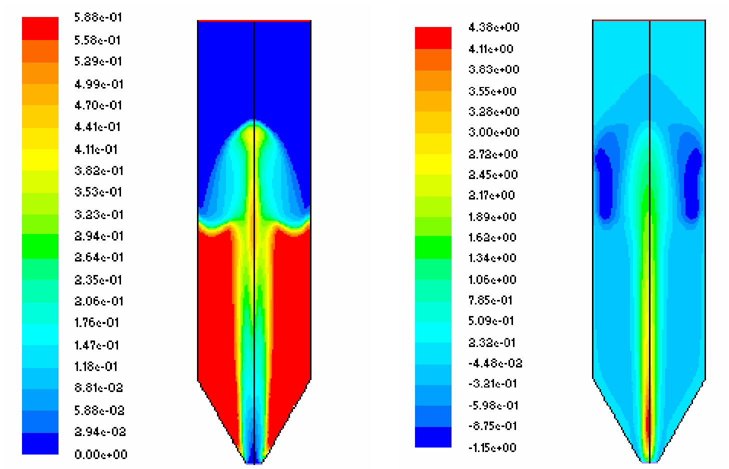
Figure 1. Scheme of equipment geometry and computational mesh.

The boundary condition was defined as velocity inlet at the entrance of the equipment, using a UDF function to represent the velocity profile in this position. An outflow boundary condition was used at the outlet, where the velocity gradient is zero. At the wall, no slip boundary condition was assumed.

4. RESULTS AND DISCUSSION

4.1. Model evaluation

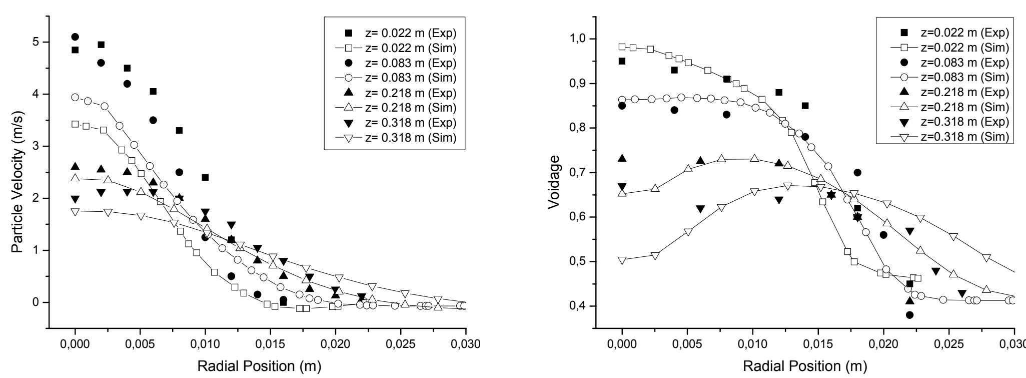
The validation of the model used to predict the behavior of multiphase flow in this work was performed by using simulated results and experimental data from He *et al.* (1994a) and He *et al.* (1994b), which used optical fiber probe measure systems. The numerical simulations were conducted using spouted bed parameters from He *et al.* (1994a). The proposed model calculated the same fountain height (15 cm) as the one experimentally verified by He *et al.* (1994a). The simulated contours of solid volume fraction and axial particle velocities can be viewed in Fig. 2.



(a) (b)

Figure 2. Results from numerical simulation: (a) contours of solid volume fraction; (b) axial particle velocities.

Figure 3 compares the simulated results with the experimental data of the solid volume fraction and axial particle velocities for many axial positions along the bed.



(a) (b)

Figure 3. Experimental and simulated profiles at different axial positions: (a) particle velocity; (b) voidage.

In Fig. 3a, near the air entrance, it can be observed a higher deviation between simulated results of particle velocities and their experimental data. It must be highlighted that at this region, the fluid behavior is very unpredictable, since the fluid stops flowing in an empty tube and starts flowing inside a bed of particles, disturbing the flow. This behavior makes the experimental measurement a difficult task to perform and the comparison with the numerical simulation is less reliable . A similar behavior for the simulated particle velocity can be observed in the simulated results of Du *et al.* (2006), although they used a different model. The radial voidage profiles present better agreement with experimental data near the fluid entrance. In this case, at this region, the fluid carries the particles axially, making the behavior of these particles more predictable due to the absence of radial components for the particle movement.

The comparison between experimental and numerical results suggests that the model presented in this work is representative of the spouted bed behavior considered.

4.2. Numerical evaluation of scale-up relations

In order to evaluate numerically the scale-up relations presented in He *et al.* (1997) for spouted beds, several simulations were performed using beds with small sizes at different scales. Looking at the scale-up relations presented in Tab. 1, it can be observed that it is not possible to satisfy the full set of relations using the same fluid. A change in fluid or particle properties must be carried out for a successful scale-up. For example, to satisfy the relation ρ_s/ρ_f, the fluid viscosity will necessarily be changed. In another way, fixing the fluid viscosity, the fluid density or solid density needs to be changed. Attention must be given to the particle diameter as well, because larger vessels will need bigger particle diameters, in order to keep the same scale factors.

All the fluid and solid properties (viscosity and density of the phases) were chosen in order to satisfy all the scale-up relations. The systems parameters, solids and fluid properties used in this analysis are presented in the Tab. 1 for three spouted beds with different sizes (SB2, SB3, SB4). The scale-up relations are also showed in the Tab. 1 as well.

Table 1. Scale factors and scale-up relations used to fulfill all the scaling procedure suggested by He *et al.* (1997).

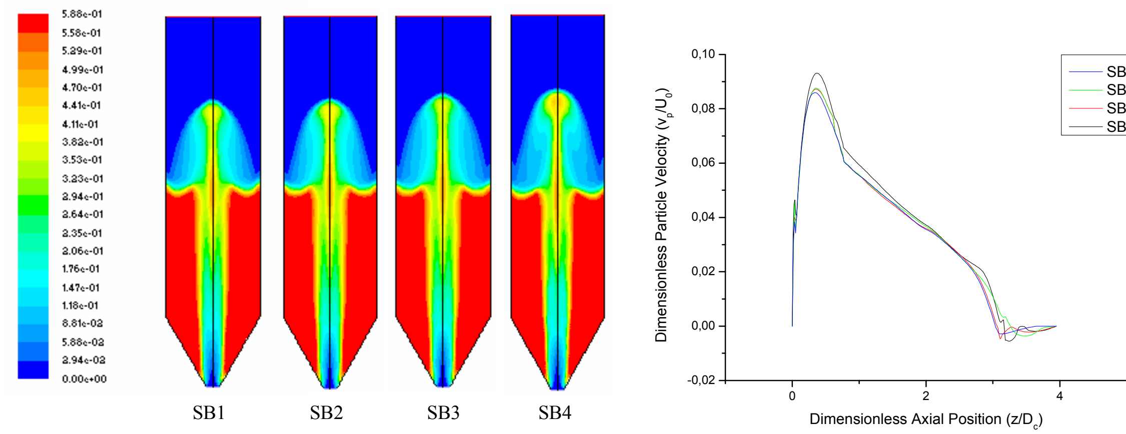
	He et al. (1994a)	***Scaled Systems***		
System	***SB1***	***SB2***	***SB3***	***SB4***
Scale Factor	***X 1***	***X 1.5***	***X 2***	***X 2***
Column diameter - D_c (m)	0.152	0.228	0.304	0.304
Inlet orifice diameter - D_i (m)	0.019	0.02865	0.0382	0.0382
Static bed height - H (m)	0.325	0.4875	0.65	0.65
Solid diameter - d_s (m)	0.00141	0.00212	0.00282	0.00282
Solid density - ρ_s (kg/m3)	2503	1362.46	2503.0	884.94
Gas density - ρ_g (kg/m3)	1.225	0.667	1.225	0.4331
Gas viscosity - μ (Pa.s)	1.810e-5	1.810e-05	5.119e-05	1.810e-05
Superficial velocity - U (m/s)	0.805	0.986	1.138	1.138
Scale-up relations				
H/Dc	2.138	2.138	2.138	2.138
D_c/D_i	7.958	7.958	7.958	7.958
D_c/d_s	107.80	107.80	107.80	107.80
ρ_s/ρ_f	2043.26	2043.26	2043.26	2043.26
E_0	0.41	0.41	0.41	0.41
Φ	1.0	1.0	1.0	1.0
Re	76.846	76.846	76.846	76.846
$U^2/g.d_s$	46.882	46.882	46.882	46.882
$\rho_s.d_s.U/\mu$	157018.6	157018.6	157018.6	157018.6
$U^2/g.D_c$	0.434	0.434	0.434	0.434

In the Tab. 1, the column "SB1" shows the parameters used by He *et al.*(1994) in their experimental study. Columns "SB2" and "SB4" show the same parameters calculated after fixing the fluid viscosity and column "SB3" presents the parameters of a scaled vessel which uses the same fluid and solid densities. In all these cases, the particle diameter also has to change together with some physical properties of the fluid in order to fulfill all scale-up relations. The scale factor is defined as:

$$Scale\ Factor = \frac{D_{ci}}{D_{c1}} \tag{14}$$

where D_{ci} is the column diameter of the scaled system and D_{c1} is the column diameter of the spouted bed of He *et al*. (1994a).

The simulated voidage contours for the systems presented in the table 2 (SB1, SB2, SB3, SB4) are presented in the Fig. 4.



(a) (b)

Figure 4. Scale-up comparison of scaled spouted beds based on the simulated: (a) voidage contours; (b) particle velocity along the spout axis (r=0)

It can be verified, in the Fig. 4a, that the behavior of solid volume fraction in all spouted beds is very similar. According to the quality of these scale-up relations, the numerical results are in agreement with the results from He *et al.* (1997), who tested these relations experimentally for small spouted beds and also obtained good results. A better visualization of the simulated results can be performed in the Fig. 4b, where it can be viewed the particle velocity behavior in the spout axis (r=0) for the dimensionless axial coordinate.

Figure 4b shows the good agreement between the particle velocity evaluated from four spouted beds with different scales. These results indicate that these relations are quite good for scale-up procedure. As He *et al.* (1997) experimentally verified, the full set of scaling relations is valid for small-scale spouted beds. By the way, in order to satisfy this full set of relations, necessarily the solid phase and the fluid phase must be changed.

4.3. Scale-up fixing solid and fluid characteristics

The scale-up relations of He *et al.* (1997) indicates that the scale-up of spouted beds from beds of small size using the same particle and fluid it cannot be achieved, since the same particle and fluid can not be used to fulfill all the relations proposed by them. Besides, the relation D_c/d_s also imposes that for a different D_c, a different d_s is necessary.

From an industrial point of view, the scale-up of spouted beds must be performed by keeping the same fluid and particle properties. Therefore, several simulations were performed in order to investigate the fluid dynamic behavior of spouted beds with similar particle diameter, fluid viscosity, fluid density and solid density. The geometrical characteristics, solid and fluid properties of the simulated beds are presented in Tab. 2.

The air velocity in the entrance was calculated from one of the scale-up relations presented by He *et al.* (1997), which correlates the air velocity in the bed with the vessel diameter. This relation is presented in Eq (15).

$$\frac{U^2}{g.D_c} \tag{15}$$

For the spouted beds SB5, SB6, SB7 and SB8, it can be observed from table 2, that some beds do not have the same scale-up values (boldface values in the lower part of table 2: scale-up relations), since d_s, ρ_s , ρ_f *and* μ were kept constant during the next scale-up simulations (boldface values in the upper part of Tab. 2: scale factors). According to He *et al.* (1997), these parameters have a strong influence in the spouted bed behavior when the system is scaled-up. In Fig. 6, simulated results for the solid volume fraction contours are presented for the scaled spouted beds indicated in Tab. 2.

Table 2. Scale factors and scale-up relations used to guarantee the scaling procedure for similar fluid and particle properties.

	He et al. (1994a)	*Scaled Systems*			
System	***SB1***	***SB5***	***SB6***	***SB7***	***SB8***
Scale Factor	***X 1***	***X 1.25***	***X 1.5***	***X 1.75***	***X 2***
Column diameter - D_c (m)	0.152	0.19	0.228	0.266	0.304
Inlet orifice diameter - D_i (m)	0.019	0.023875	0.02865	0.033425	0.0382
Static bed height - H (m)	0.325	0.40625	0.4875	0.56875	0.65
Solid diameter - d_s (m)	0.00141	**0.00141**	**0.00141**	**0.00141**	**0.00141**
Solid density - ρ_s (kg/m3)	2503	**2503**	**2503**	**2503**	**2503**
Gas density - ρ_g (kg/m3)	1.225	**1.225**	**1.225**	**1.225**	**1.225**
Gas viscosity - μ (Pa.s)	1.810e-5	**1.810e-05**	**1.810e-05**	**1.810e-05**	**1.810e-05**
Superficial velocity - U (m/s)	0.805	0.900	0.986	1.065	1.138
Scale-up relations					
H/Dc	2.138	2.138	2.138	2.138	2.138
D_c/D_i	7.958	7.958	7.958	7.958	7.958
D_c/d_s	107.80	**134.75**	**161.70**	**188.65**	**215.60**
ρ_s/ρ_f	2043.26	2043.26	2043.26	2043.26	2043.26
E_0	0.41	0.41	0.41	0.41	0.41
Φ	1.0	1	1	1	1
Re	76.846	**85.917**	**94.117**	**101.658**	**108.677**
$U^2/g.d_s$	46.882	**58.6032**	**70.323**	**82.044**	**93.765**
$\rho_s.D_p.U/\mu$	157018.6	**175552.1**	**192307.7**	**207716.1**	**222057.8**
$U^2/g.d_s$	0.434	0.434	0.434	0.434	0.434

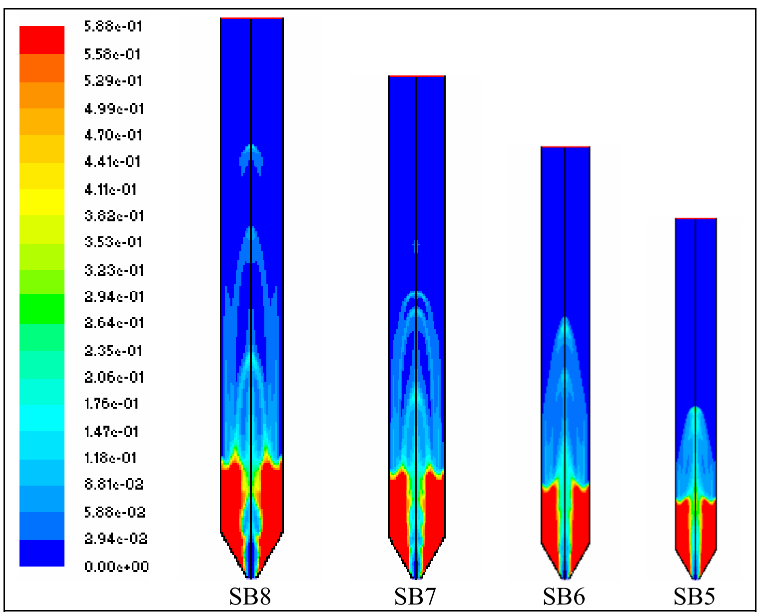
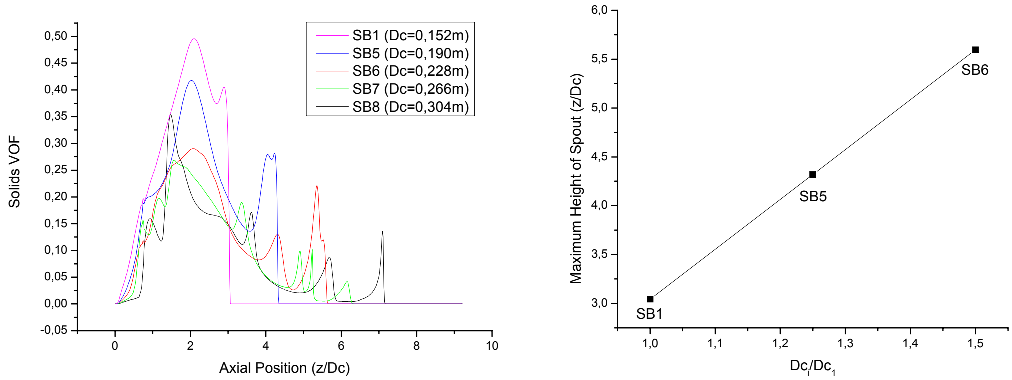


Figure 6 – Contours of solids volume fraction for different scales with same solid and fluid properties.

Spouted beds with scaled up to factors of 1.5 or less (SB5, SB6) presented stable spouts, with a well-defined fountain and annulus regions, and of course the spout. However, spouted beds with scale-up factors of higher than 1.5 presented a pulsed fountain, which characterizes an unstable spouted bed behavior. Figure 7a presents several peaks of solid volume fraction, at high heights, for spouted beds SB6, SB7 and SB8, which indicates a non constant fountain. This unstable operation is due to an overestimated air velocity at the inlet, which was calculated by Eq. (15) for scaled vessels. Figure 7b shows the axial solid volume fraction profiles along the axis of the spouted bed and the fountain height for the stable beds. In the stable scaled spouted beds (SB5, SB6), it is possible to verify a linear relationship between the maximum height of the fountain and the scale factor, indicating a predictable behavior.



(a) (b)

Figure 7. (a) Solid volume fraction behavior in the axis of the spouted beds; (b) Maximum height of spout for stable spouted beds.

These pulses verified in the results on spouted beds for high air flow rates were already experimentally verified (Xu *et al.*, 2004). This behavior indicates the presence of a maximum scale factor, where the limit of stability is reached near to 1.5 for the scale factor. High scale factors applied to these systems are not recommended, due to the instability observed in the spouted beds simulated after the overestimation of the inlet velocity, when the scaled-up relation of Eq. (15) is kept constant.

5. CONCLUSION

A numerical study of the scale-up for spouted beds was performed in this work. It was verified that the scale-up relations of He *et al.* (1997) are quite consistent and they are in agreement with the experimental results obtained for small vessels by the authors. A scale-up procedure based on these relations is not possible if the same fluid is used during scale-up. A different fluid must be used in order to satisfy the full set of scale-up relations suggested by He *et al.* (1997). The study of scale-up procedure for constant particle and fluid characteristics indicated an overestimated inlet velocity, calculated from the scale-up relations of He *et al.* (1997). A linear behavior between the fountain height and the scale-up factor was observed, and it must be investigated in the future.

6. ACKNOWLEDGES

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