Engineering Conferences International
ECI Digital Archives

The 14th International Conference on Fluidization – From Fundamentals to Products

Refereed Proceedings

2013

Estimation of the Circulating Time in a Largescale Fluidized Bed Using DBM Simulation Data

Javier Sánchez-Prieto
Universidad Carlos III de Madrid, Spain

Sergio Sánchez-Delgado
Universidad Carlos III de Madrid, Spain

Néstor García Hernando
Universidad Carlos III de Madrid, Spain

Javier Villa Briongos
Universidad Carlos III de Madrid, Spain

ESTIMATION OF THE CIRCULATING TIME IN A LARGE-SCALE FLUIDIZED BED USING DBM SIMULATION DATA

Javier Sánchez-Prieto[a]*, Sergio Sánchez-Delgado[a], Néstor García Hernando[a] and Javier Villa Briongos[a]
[a]Universidad Carlos III de Madrid, Thermal and Fluid Engineering Department. Avda. de la Universidad 30, 28911 Leganés (Madrid), Spain
*T: +34916246032; E: jsprieto@ing.uc3m.es

ABSTRACT

In this work, the Discrete Bubble Model (DBM) reported by Briongos *et al.* (1) is used to calculate bubble phase data in a large-scale bubbling fluidized bed. The data provided by the model was used to calculate the circulation time, as the time required for a group of particles to reach the freeboard from the bottom of a fluidized bed and return to their original height, by means of a new methodology proposed by Sánchez-Delgado *et al.* (2). The results are consistent with experimental works found in the literature, showing similar trends for the lab-scale bed and for the scale-up of this one.

INTRODUCTION

Bubbling gas-solid fluidized beds (FB) are broadly applied in industry, particularly in thermochemical energy conversion processes such as combustion and gasification. In this kind of processes the movement of solids, mixing and particle segregation are very important features for modeling and scale-up of fluidized bed reactors. The fluidization process offers a high heat transfer rate, good gas-solid mixing and solid handling, and provides a uniform and controllable temperature. Because of all this advantages, fluidization is a very promising technology for the valorization of biomass and wastes in energy conversion processes. The high complexity characterizing conventional gas-solid FBs dynamics increases when dealing with biomass due to the limited research reported.

To properly design a process in a fluidized bed, a comparison between the circulation time of the reactor and the characteristic reaction time for a specific application is needed. This comparison is also useful for the modelling of fluidized beds to verify the assumption made in the model for different purposes (countercurrent, back-mixing, well-mixing, plug-flow). To ensure that the fluidized bed is well mixed, the characteristic circulation time of the specific application needs to be similar to the circulation time. Besides, the circulation time obtained with the estimation can be useful to validate Dynamic Numerical Simulation models.

Numerical modeling of these systems has significantly been developed over the last two decades. The FB modeling can be divided into two main groups:

Eulerian-Eulerian models that consider the gas and solid phases as interpenetrating continua, and Eulerian-Lagrangian models that couple a Lagrangian description of the particle dynamics with a continuum description of the gas phase, such as discrete particle model (DPM) or discrete bubble model (DBM). The DPM approach tries to describe the large-scale dynamics of FBs by direct modeling of the particle-particle and gas-particle interactions. Although DPM is a very powerful tool to characterize the multiscale flow structure of FBs, the computational time needed to model a large number of particles limits it use when modeling large-scale FBs. DBM models, as reported by Pannala *et al.* (3) and Bokkers *et al.* (4), focus on the bubble interaction and have shown to be useful for modeling global dynamics of large-scale FBs. The interacting Discrete Bubble Model (DBM) approaches reported in literature have show to be a fast and useful tool for modeling global dynamics of large-scale fluidized bed with low computational cost. Despite of this important advantages, DBM models only provide information about bubble phase dynamics.

The circulation time is defined as the time required for a group of particles to reach the freeboard from the bottom of a fluidized bed and return to their original height. This parameter can be estimated with a new methodology proposed by Sánchez-Delgado *et al.* (2) based on the calculation of the turnover time defined by Geldart (5) and the visible bubble flow reported by Grace y Clift (6). Using this methodology, information about the bed dense phase can be obtained from bubble phase data.

In this work the DBM approach reported by Briongos *et al.* (1), coupled with the methodology developed by Sánchez-Delgado *et al.* (2), is used to estimate the solids circulation time in a large-scale bubbling fluidized bed.

SIMULATION

As stated above, with the use of DBM code reported by Briongos et al. (1), two different beds were simulated. Firstly the experimental facility described in Sobrino *et al.* (7) and secondly a scaled-up version of this one. The main design and operation conditions of the two facilities are shown in Figure 1.

	Lab-scale	**Large-scale**
H_b [m]	0.22	3.52
D_t [m]	0.193	3.08
U_0 [m/s]	0.57	0.57
U_{mf} [m/s]	0.36	0.36
d_p [μm]	540	540
ρ_p [kg/m^3]	2632.5	2632.5
ε_0	0.455	0.455
Distributor type	Multiorifice	Tuyere
Number of orifices	90	-
Number of tuyeres	-	286
Orifices per tuyere	-	20
Orifice diameter [mm]	2	4
Simulation time [s]	600	240

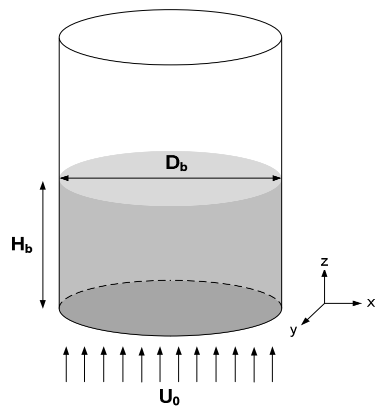

Figure 1: Scheme of the simulated bed and main parameters selected for the simulation.

The scale-up was developed remaining constant the geometry, the bed aspect ratio and the ratio between the area of each orifice in the distributor and the bed cross-sectional area. The multi orifice of distributor was changed to tuyere type to reduce the computational cost.

Taking into account the activation region mechanism (1), bubbles below 0.05 m height were removed. The first 30 seconds of simulation were also removed to ensure that the simulation is in the steady state. After this filtering process, the bubble phase data was used to calculate the circulation time.

METHODOLOGY

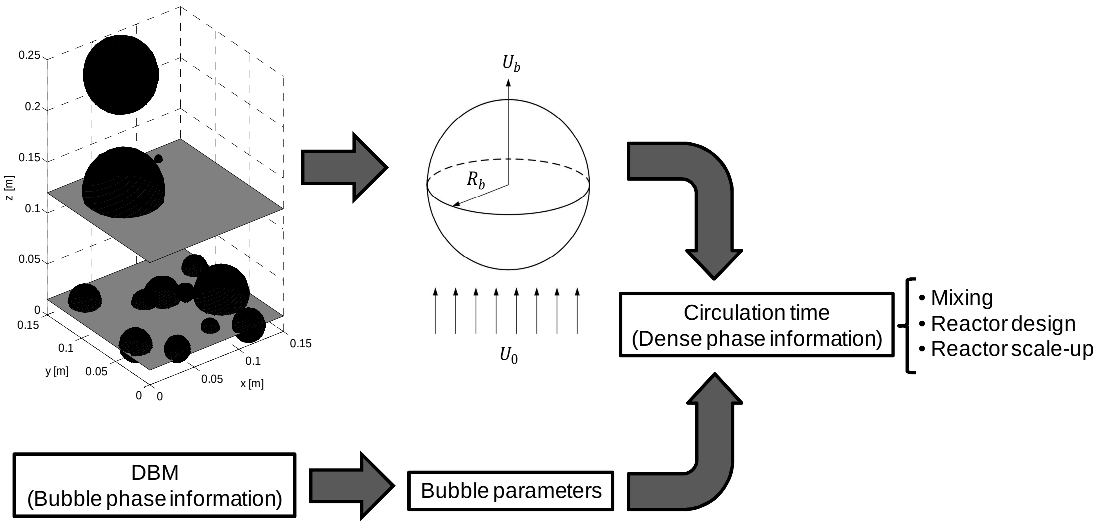

Figure 2: Scheme of the methodology for the estimation of circulation time using DBM simulation data.

The circulation time is defined as the time required for a group of particles to reach the freeboard region from the bottom of the fluidized bed and return to the original height. The circulation time can be calculated using equation [1], that is based on the turnover time (t_T) reported by Geldart (5).

$$t_{c,est} = 2t_T = 2\frac{M}{M'} \qquad [1]$$

Where M is the total mass of solids in the bed, and M' is the mass flow rate of solids. This equation can be expressed as a function of the bulk density, ρ_p', the cross-sectional area of the bed, A, the visible bubble flow, Q_b, and the effective freeboard height, h' (Equation [2]). The effective freeboard height, h', is defined as a function of the free board height and the minimum height where bubbles generation takes place, $h' = h_{fb} - h_{min}$. In the DBM, bubbles are generated using the activation region approach and, therefore h_{min} was selected as the activation region height (h_{min} = 0.05 m). The freeboard height, h_{fb} (mean distance from the distributor to the top of the bed during fluidization), was obtained experimentally by Sánchez-Delgado *et al.* (2) using DIA techniques. The authors found values of h_{fb} between 1.1-1.2 times the fixed bed height. According to that, in the DBM the freeboard height was selected to be 1.15 times the fixed bed height.

$$t_{c,est} = 2\frac{\rho_p' A h'}{\rho_p' Q_b} = 2\frac{Ah'}{Q_b} \qquad [2]$$

The circulation time estimated with equation [2] only depends on operating parameters (A and h') and the visible bubble flow (Q_b), calculated with the bed thickness, bubble diameter, D_b, and the bubble velocity, U_b (Equation [3]).

$$Q_b = \frac{1}{N}\sum_{j=1}^{N}\sum_{i=1}^{n} U_{bi} a_i \qquad [3]$$

Where a_i is the area of the i^{th} bubble cut by a horizontal section (Figure 2), n is the number of bubbles passing through this section at i^{th} frame, j is the current frame number (from $j=1$ to $j=N$) and N the total number of images analyzed.

A schematic picture of the whole process is presented in Figure 2.

RESULTS AND DISCUSSION

To calculate the visible bubble flow as is defined in equation [3], a proper section at a given height of the bed has to be selected. The circulation time was estimated for the lab-scale unit (Figure 1) at different section heights, h_{cut}.

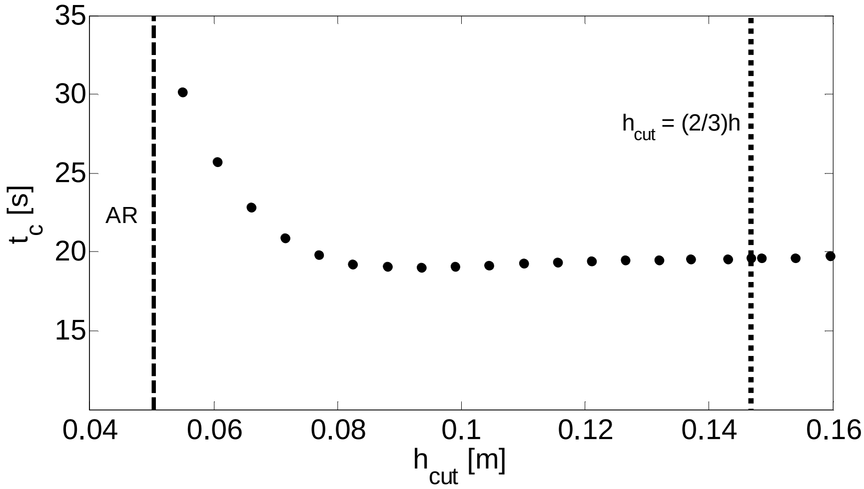

Figure 3: Circulation time at different section heights for the lab-scale unit.

As can be observed in figure 3, the circulation time is overestimated if the section is taken close to the distributor, due to the presence of a huge amount of small and slow bubbles before that the coalescence effect take place in the bed. Larger bubbles generated by coalescence are the drivers of the dynamics of the bed. This overestimation decreases as long as the section height increases reaching a constant value. Also, if this section is very close to the freeboard region the estimation can be affected by the splash zone. Taking it into account, the section height for the calculations was selected to be 2/3 times the fixed bed height.

The circulation time was calculated for both lab and large-scale beds, at different excess air-ratio.

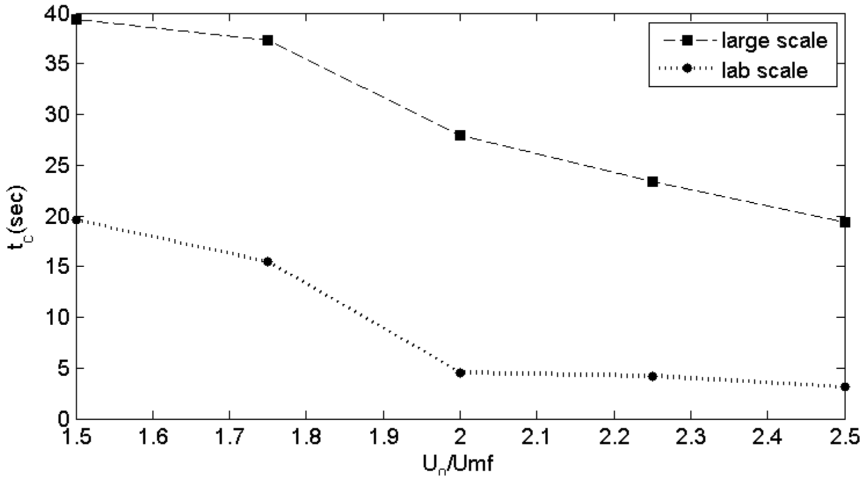

Figure 4: Circulation time at different gas velocities for both lab-scale and large-scale beds.

As expected, for both cases, the circulation time decreases when the gas velocity increases. At higher velocities, the solid mixing is enhanced and therefore the circulation time decreases. Similar trends were found in both lab and large-scale beds, that suggests that circulation time is strongly related to the bed geometry.

The results of the lab-scale bed were compared to the experimental results of (2). They found an asymptotic trend in the circulation time in 2D beds of different heights. As can be seen in figure 4, the DBM is able to reproduce this trend in both lab and large-scale beds. Simulations of the lab-scale at a different height were run to show if this effect is present in 3D beds.

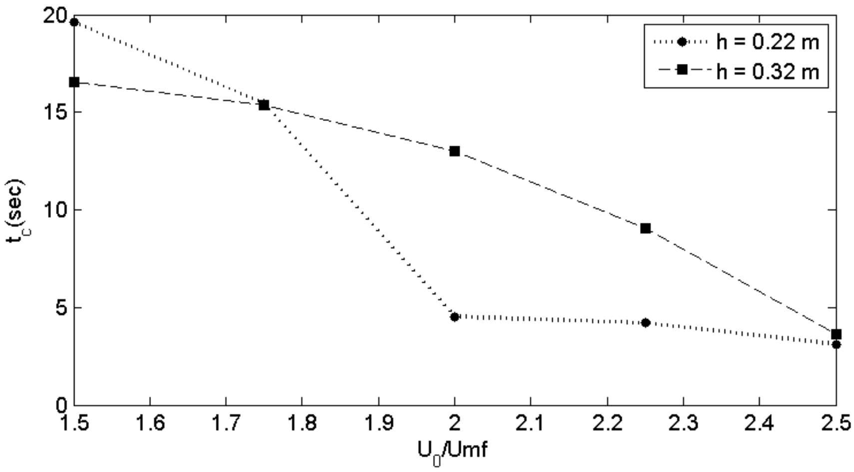

Figure 5: Circulation time at different gas velocities for different bed heights.

At a higher bed height the aspect ratio of the bed changes and also the variation of the circulation time with the gas velocity. The circulation time at low gas velocity is smaller for the 0.32 m height bed than for the 0.22 m height. This is contrary to that can be expected because solids need more time to reach a higher height. This fact suggests that there are other factors affecting to the circulation time such the aspect ratio of the bed.

In this work, the circulation time has an asymptotic trend when the superficial gas velocity increases. This effect is directly related with the work of Cui et al. (8) and Cui et al. (9), where an expression for the bubble fraction as a function of the superficial gas velocity is presented. When the superficial gas velocity is increasing, the bubble fraction also increases until a certain value where the bubble fraction keeps constant. At this moment, if the superficial gas velocity increases, the excess gas has no effect on the bubble fraction, but produces a relevant increase of the throughflow. Therefore, this limitation of the bubble fraction is also a limitation of the circulation time with the superficial gas velocity in a bubbling fluidized bed.

CONCLUSIONS

The solid circulation time in a fluidized bed can be predicted with combination of DBM simulation data and a novel methodology based on the turn over time concept. Two different beds were simulated, a lab-scale fluidized bed and a scaled-up version of this one. Similar trends were found for the circulation time with increasing gas velocity for both simulated beds. Using the same aspect ratio, even for larger beds, the circulation time follows the same trend with velocity. At different fixed bed heights, this trend changes, probably affected by the aspect ratio of the bed.

The simulations were compared to experimental results found in literature for 2D beds. A similar asymptotic behavior was found for the circulation time, reaching a low value at higher velocities. This behavior was also found in beds with different aspect ratios.

DBM simulation coupled with the methodology described to estimate the circulation time is a fast and useful tool to gather information about mixing in a large-scale fluidized bed.

In this work, the validated technique to calculate the circulation time in a 2D fluidized bed, has been applied in a 3D simulated bed by DBM approach. Therefore, as a future work, the circulation time results obtained in the 3D simulation should be compared with 3D experimental results (X-ray, magnetic resonance, objects movement in the bed surface, optical probes...). Besides, experimental result of a large scale facility can be useful in the scale-up process, however, these results are not always available

ACKNOWLEDGEMENTS

The financial support from projects DPI2009-10518 (MICINN) and CARDENER-CM (S2009ENE-1660) is very much acknowledged.

NOTATION

a_i	Area of the i^{th} bubble cut by a horizontal section [m^2]
A	Cross-sectional area of the bed [m^2]
d_p	Particle diameter [µm]
D_t	Bed diameter [m]
D_b	Bubble diameter [m]
h	Fixed bed height [m]
h'	Effective height [m]
h_{fb}	Height of the freeboard [m]
h_{min}	Activation region height [m]
M	Mass of solid particles in the fluidized bed [kg]
M'	Mass flow rate [kg/s]
N	Total number of images [-]
Q_b	Visible bubble flow [m^3/s]
t_c	Circulation time [s]
t_T	Turnover time [s]
U_0	Superficial gas velocity [m/s]
U_b	Bubble velocity [m/s]
U_{mf}	Minimum fluidization velocity [m/s]
ρ_p'	Particle density [kg/m^3]
ε_0	Fixed bed voidage [-]

REFERENCES

1. J.V. Briongos, S. Sánchez-Delgado, A. Acosta-Iborra, D. Santana. A novel approach for modeling bubbling gas-solid fluidized beds. AIChE J., 51:734-744, 2010.
2. S. Sánchez-Delgado, C. Marugán-Cruz, A. Soria-Verdugo and D. Santana. Estimation and experimental validation of the circulation time in a 2D gas-solid fluidized bed. Powder Technology (2012), doi: 10.1016/j.powtec.2012.11.012.
3. Pannala, S., Daw, CS., Hallow, J.S. Dynamic interacting bubble simulation (DIBS): an agent-based bubble model for reacting fluidized beds. Chaos, 14: 487-498, 2004.
4. Bokkers, G.A., Laverman, J.A., van Sint Annaland, M., Kuipers, J.A.M. Modeling of a large-scale dense gas-solid bubbling fluidized bed using a novel discrete bubble model. Chem. Eng. Sci. 61:5590-5602, 2006.
5. D. Geldart. Gas fluidization technology. Wiley-Interscience Publication, 1986.
6. J. Grace and R. Clift. On the two-phase theory of fluidization. Chem. Eng. Sci., 29:327-334, 1974.
7. Sobrino, C., Acosta-Iborra, A., Santana, D., de Vega, M. Bubble characteristics in a bubbling fluidized bed with a rotating distributor. International Journal of Multiphase Flow, 35:970-976, 2009.
8. H. Cui, N. Mostoufi, J. Chaouki, Characterization of dynamic gas–solid distribution in fluidized beds. Chemical Engineering Journal, 79:135-143, 2000.
9. H. Cui, N. Mostoufi, J. Chaouki, Gas and solids between dynamic bubble and emulsion in gas-solid fluidized beds. Powder Technology, 120:12-20 2000.