Refereed Proceedings

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering

Engineering Conferences International *Year* 2007

Experimental Analysis of Fuel Mixing Patterns in a Fluidized Bed

David Pallarès* Pedro A. Díez†
Filip Johnsson‡

*Chalmers University of Technology, david.pallares@chalmers.se
†École Nationale Supérieure de Techniques Avancées, diezbrea@ensta.fr
‡Chalmers University of Technology, fijo@chalmers.se

EXPERIMENTAL ANALYSIS OF FUEL MIXING PATTERNS IN A FLUIDIZED BED

David Pallarès, Pedro A. Díez, Filip Johnsson
Department of Energy and Environment
Chalmers University of Technology
SE-41296 Göteborg (Sweden)
T: +46-317721449; E: filip.johnsson@me.chalmers.se

ABSTRACT

The mixing pattern of a tracer particle which simulates a fuel particle is studied in a cold 2-dimensional fluidized bed with respect to the influence of fluidization velocity, bed height, tracer particle size and air-distributor pressure drop under conditions typical for bubbling fluidized bed boilers as well as the bottom region of circulating fluidized bed boilers. The results show that for all conditions studied, the tracer particle follows a flow pattern structured into horizontally-aligned vortexes.

INTRODUCTION

Fuel mixing is a key phenomenon for the performance of fluidized bed (FB) boilers (bubbling as well as circulating). In the vertical direction, good fuel mixing is important to ensure sufficient contact time between fuel and combustion air. In the horizontal direction, the fuel dispersion determines to what extent there will be homogeneous cross-sectional fuel distribution which is important for fuel burnout while allowing a low excess air ratio (thus minimizing operational costs). For typical fluidization conditions, horizontal solids mixing in the bottom region of an FB unit was found to be lower than in the vertical direction (1). This, together with the fact that the bed of an FB boiler (where most of the fuel inventory is present) usually has a height-to-width ratio lower than 1, makes the horizontal direction critical in terms of fuel mixing. Moreover, for economical reasons, the number of fuel feeding points should be kept as low as possible (2), which is obviously strongly related to the degree of horizontal fuel mixing.

To what extent a certain fuel mixing behaviour is sufficient or not depends on the fuel conversion time and the characteristic mixing length. A comparison of the characteristic times for fuel dispersion and conversion can be expressed by the Damköhler number:

$$Da = \frac{\tau_{dispersion}}{\tau_{conversion}} = \frac{L^* / r_{dispersion}}{\tau_{conversion}} \qquad (1)$$

Thus, the Damköhler number (*Da*) is a suitable parameter for evaluation of fuel mixing in FB units. The *Da* number indicates whether the dispersion rate is high enough to ensure a sufficiently homogeneous distribution of the fuel over the entire cross section of the unit (which is the case for low values of the Damköhler number, *Da*<1). It is seen from Eq. (1) that operational conditions which yield a sufficient fuel mixing rate in a certain FB burning a certain fuel may not be sufficient when changing fuel (*e.g.* to a fuel with a higher volatile content or which is more reactive). It is evident that horizontal fuel mixing becomes a critical issue in large CFB boilers, which may have a cross-sectional area of up to several hundreds of square meters.

To the authors knowledge there are no models on the fuel mixing which give satisfactory agreement with experimental data under conditions applicable to FB boilers and which are based on the underlying physics of the mixing process. Although, there has been significant progress in numerical modelling from first principles, computational fluid dynamics (CFD) is generally limited to a mono-sized solid phase (3-5), while simulating fuel mixing obviously requires accounting for a polydispersed solid phase (bulk particles and fuel particles). There are some attempts in literature to account for polydispersed solids in CFD simulations (6, 7) but not really concerning fuel mixing, which requires one of the phases (fuel) to have a much smaller fraction (less than 5%) than that of the bulk and to be lighter as well. An attempt to implement conditions corresponding to fuel mixing was made by Tanskanen (8), but more work is required until realistic results can be obtained. Thus, there is a need for semiempirical models expressing the solids mixing as an overall dispersion coefficient to be used as a tool to simulate horizontal fuel dispersion (the word “dispersion” is here used for simplicity since measured and modelled solids mixing is normally expressed as an average dispersion coefficient, although the mixing process is highly convective). Effective horizontal solids dispersion coefficients in FB boilers have been estimated by means of experimental data from boilers (9, 10) and from cold rigs (11). Niklasson *et al*. (9) carried out experiments in the Chalmers 12 MW FB boiler operated under bubbling conditions, obtaining a value of the horizontal fuel dispersion coefficient of around 0.1 m^2/s, a result which seems to be consistent with the boiler values reported by Xiang *et al*. (10), ranging from 0.01 to 0.1 m/s^2 for fluidization velocities lower than those applied in (9). However, experimental evaluation of the dispersion only in the form of global dispersion coefficients is a limitation in the sense that this gives no information on the mechanisms governing the mixing process. Application of particle tracking techniques can provide an experimental basis for resolving the particle mixing process in both space and time.

Several particle tracking techniques have been used to investigate solids mixing in FB units (*cf.* 12) for a survey of experimental investigations on fluidized-bed solids mixing). Techniques measuring in a fixed point (Eulerian) exist, such as Laser Doppler Anemometry or the use of radio transmitters, but no information on the trajectory of a single particle can be obtained from such techniques. For tracking in 3-dimensional units, various tomographical techniques have been developed based on X-ray, electrical capacitance or γ-ray emission. The latter was applied by Stein *et al.* (13) in scale models of FB boilers under conditions accounting for fluiddynamical scaling relationships. Experiments in 2-dimensional rigs have visual access to the dynamics also in dense beds as the main advantage but are obviously limited to qualitative studies. Most experimental works focus on studying the mixing of the bed material itself, *i.e.* using a tracer particle to mimic the bulk bed material. Accordingly, a tracer particle with larger size and lower density than those forming the bulk bed material must be applied when mixing of a fuel particle is studied, as done in (13-16). Yet, none of these works gives the velocity and concentration fields of the tracer particle while varying main operational parameters. The authors of the present work studied the mixing pattern of a tracer particle simulating a fuel particle in a cold 2-dimensional fluidized with the wide dimension being 0.4 m (12), finding the flow pattern of the tracer particle to be structured into several horizontally-aligned vortexes with alternated rotational direction. The question is to what extent this is also valid in a wider unit since the limited width of the bed (0.4m) applied in the previous work may have influenced the horizontal spreading of the tracer particle. Thus, the present work extends the previous work with the aim to further generalize the patterns of the fuel mixing process, with focus on operational conditions typical for fluidized-bed boilers and with the experiments carried out in a 2D bed with the wide dimension about three times (1.2 m) that applied in the previous work.

EXPERIMENTAL SETUP

This work applies the particle tracking technique developed in the previous work (12), which is suitable for tracking particles coarser than the bed material (*i.e.* simulating fuel particles) in cold 2-dimensional fluidized beds and is robust in that the dynamics of the mixing can be studied over a wide range of operational parameters, allowing for a fundamental study on the phenomenology of solids mixing. The particle tracking technique is based on tracking a single tracer particle (a plastic capsule filled with a phosphorescent solution) in a 2-dimensional fluidized bed with a transparent front wall. The mixing process is then analysed and quantified by means of digital image analysis of the trajectory of the capsule. To maximize phosphorescence, the riser is placed in a dark chamber. A special, high-gain CCTV video camera with a time resolution of 4×10^{-2} s is used for filming the capsule in the bed. In addition, glass beads are used as bed material, allowing phosphorescence to be seen through the bed material as when the tracer particle flows close to the rear wall (opposite camera position). The 2-dimensional unit is illustrated in Figure 1. The riser is 1.2 m wide with a depth of 0.02 m and a height of 2.05 m with a Plexiglas front wall. The gas flow is controlled by a valve located close to the air plenum and the externally recirculated solids flow is refed into the riser through the back wall (see item no. 9 in Figure 1). Two different perforated air-distributor plates are used, both with 2 mm i.d. holes and hole areas of 2% and 9% (called "high-ΔP" and "low-ΔP" air distributor, respectively), which yields the ΔP vs u curves given in Figure 2. The glass beads forming the bed have a narrow particle size distribution with an average size of 330 µm and a density of 2500 kg/m^3, *i.e.* although no exact scaling is performed, these values are similar to those of sand or ash particles typically used as bed material in fluidized bed boilers. The glass beads belong to Group B in the Geldart classification with u_{mf}=0.12 m/s and u_t=1.76 m/s (ambient conditions). For the conditions applied, around 60 minutes of video recording of the tracer particle (providing approximately 90,000 video frames) were shown to give high enough statistical significance in the digital image analysis. Thus, the method requires significant number of data to reach statistical significance.

Table 1 lists the operational conditions for the 9 experimental runs carried out as basis for this work. The runs cover flow conditions qualitatively resembling fluidization regimes typically found in a fluidized-bed boiler. Two sizes of tracer particle were applied both with a size representative for fuels in FB boilers: 15 mm x 6 mm i.d. ("Large") and 10 mm x 4 mm i.d. ("Small"). Although the bed is only 20 mm in depth, analysis of the trajectories of the tracer particles do not indicate any significant influence of wall effects other than the above-mentioned general wall effect which, of course, makes the results qualitative. Both tracers have an apparent particle density of around 650 kg/m3, which is close to that of biomass, somewhat higher than that of coal, and lower than that of waste pellets. The different regimes were reached by changing either operational conditions (fluidization velocity, amount of bed material and tracer particle) or the air distributor.

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering, Art. 114 [2007]

Table 1: Test matrix.

	RUN								
	A	B	C	D	E	F	G	H	I
Fluidization velocity [m/s]	0.7	1.5	2.7	3.5	2.7	2.7	2.7	2.7	3.5
Fixed bed height [m]	0.33	0.33	0.33	0.33	0.33	0.18	0.18	0.18	0.33
Tracer size	Large	Large	Large	Large	Small	Large	Large	Small	Large
Air distributor	High-ΔP	High-ΔP	High-ΔP	High-ΔP	High-ΔP	High-ΔP	Low-ΔP	High-ΔP	Low-ΔP

Since the evaluation method is based on tracking a single particle in each run, interactions between fuel particles are assumed negligible. This should be a reasonable assumption since the fuel fraction usually represents only between 1 and 5% of the total bed material in FB boilers. With this assumption, the probability of presence of the tracer particle at some location can be interpreted as the concentration value at steady state, C_{fuel}, normalized so that:

$$\int_A C_{fuel} \cdot dA = 1 \tag{2}$$

This can be expressed in discrete form, more suitable for the digital image analysis:

$$\sum_{1}^{N} C_{fuel,n} \cdot A_n = 1 \qquad \text{with } n=1\ldots N \tag{3}$$

where N corresponds to the total number of pixels of the analyzed videoframe. It should be noted that the probability of not capturing any phosphorescence from the tracer particle at a certain videoframe (around 30% in the present work) is much higher when the tracer particle is immersed in the dense bottom bed than when it is flowing in the freeboard. Thus, this uneven spatial distribution of the phosphorescence capture probability would bias the concentration plots giving the bottom bed region lower concentration values. However, the location of the tracer during the short time intervals (fractions of a second) in which its phosphorescence is not captured can be estimated by linear interpolation. Although the method is restricted to 2D FB units and thereby to qualitative evaluation of the results obtained, tracer particle dispersion is quantified to enable direct comparison of the runs. Thus, the local dispersion coefficient in position n (*i.e.* pixel) in the direction k (horizontal or vertical in our case) at any location can be calculated as:

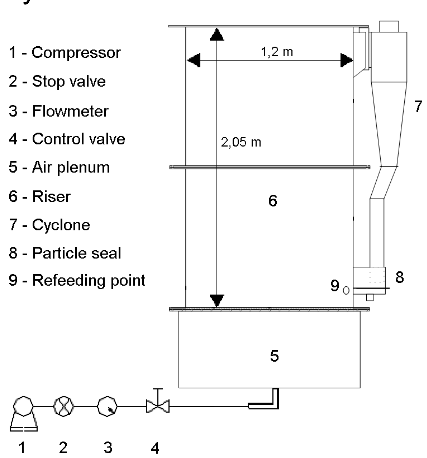

Figure 1: The cold 2-dimensional FB unit used in the experiments.

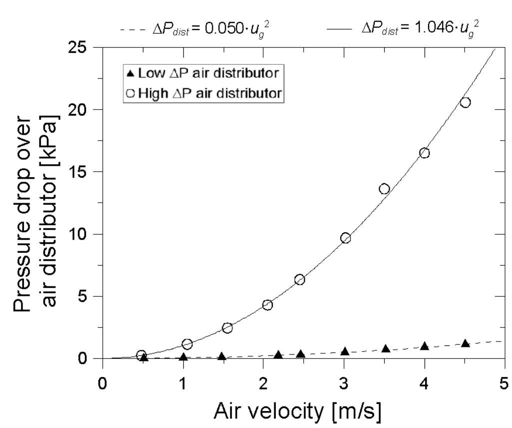

Figure 2: Characteristic curves of the air distributor plates used in the experiments.

Table 2: Average dispersion coefficients as obtained in this work.

	RUN								
	A	B	C	D	E	F	G	H	I
$\overline{D}_x$ [m²/s]·10²	0.31	1.11	1.59	2.30	1.48	1.16	1.08	1.20	4.24
$\overline{D}_y$ [m²/s]·10²	0.11	0.53	1.67	4.01	2.32	1.14	0.90	1.22	3.00

$$D_{k,n} = \frac{\Delta l_{k,n}^2}{2 \cdot \Delta t_n}; \qquad \text{with } k = x, y \quad \text{and} \quad \text{for } n = 1 \ldots N \qquad (4)$$

As discussed in (12), setting Δt_n equal to a low value and thus taking $\Delta l_{k,n}$ as the displacement occurred in that time interval would imply values of the dispersion coefficient that do not account for the history of the trajectory (which is a prerequisite for defining a dispersion coefficient) and a pseudo-velocity value would be obtained instead. Setting Δt_n to a large value to avoid this effect would increase the risk of having Δt_n larger than the characteristic time for the solids mixing in the unit, also resulting in unrealistic values of the dispersion coefficients. Instead, the method used is to set $\Delta l_{k,n}$ to a relatively high value in order to capture the history of the mixing process, which is shown to be a more robust method for application of Eq (4). Thus, using this approach, Δt_n is defined as the time it takes for the tracer particle to flow a distance equal to $\Delta l_{k,n}$, which has been set to 0.3 m in this work (a fourth of the riser width, as in (12)).

Finally, a global dispersion coefficient can be calculated by weighting the local dispersion coefficients with respect to concentration values, *i.e.*

$$\overline{D_k} = \sum_1^N C_{fuel,n} \cdot A_n \cdot D_{k,n} \qquad \text{with } k = x, y \quad \text{and} \quad \text{for } n = 1 \ldots N \qquad (5)$$

RESULTS

Table 2 lists the average dispersion coefficients obtained for the runs given in Table 1. It should be kept in mind that evaluation of dispersion coefficients in eminently convective flows (as in the present work) has to be done with care, since the values are strongly dependent on the calculation procedure ($\Delta l_{k,n}$ and Equations (4) and (5)), the 2D bed flow and, as indicated above, the convective nature of the flow. Thus, the values in Table 2 have a restricted use only for comparison between runs in this work and shall not be used in, for example, simulations based on a Brownian diffusion process. The calculation procedure (inclusion of a concentration-weighting term in Equation (5) instead of a spatial average) is the reason why the results show vertical and horizontal mixing coefficients in the same order of magnitude (in contrast to (1)). Yet, fuel mixing in FB boilers is critical in the horizontal direction due to the low aspect ratio of the dense bed (where most fuel particles are found). Removing the concentration-weighting term in Equation (5) would significantly increase the values of vertical dispersion coefficient.

When it comes to the appearance of the concentration and velocity fields, the vortex structure of the tracer flow observed in the previous work (in the 0.40 m wide bed) is also found in the present work (1.2 m wide bed) in all runs (A to I), independent of fluidization regime. Thus, it seems to be a general feature that the flow of a tracer particle of the size and density similar to a fuel particle is structured into several horizontally-aligned vortexes with alternated rotational direction (*cf.* 12). Figure 3 exemplifies this for Run C, with the ascending channels provided by the bubble paths located at the approximate coordinates *x*=0.20 m and *x*=1.0 m. Thus, the vortex structure is induced by the bubble flow rising through the dense bottom bed, which provides ascending channels for the tracer particle through the dense bottom bed and up to the bottom bed surface, where the tracer particle is projected into the splash zone (experiencing a horizontal displacement) before sinking in the bottom bed through the emulsion phase. This 2D vortex flow structure can be seen as a vertical cross section of what would be a toroidal flow structure around each bubble path in a 3-dimensional case.

An increase in fluidization velocity clearly has a beneficial effect to the solids mixing as can be seen in Figure 4, where the average dispersion coefficient increases in both the vertical and horizontal direction with an increase in velocity, mainly due to the increased bubble flow.

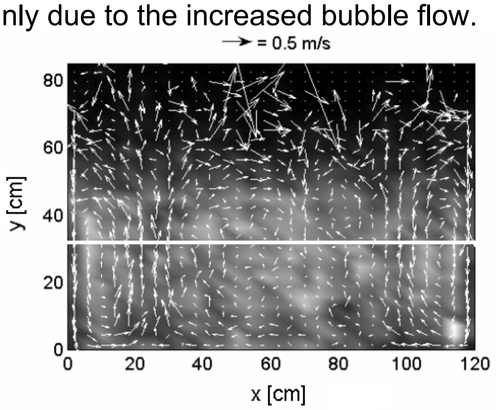

Figure 3: Tracer particle concentration and velocity fields for Run C. White line indicates the bed surface.

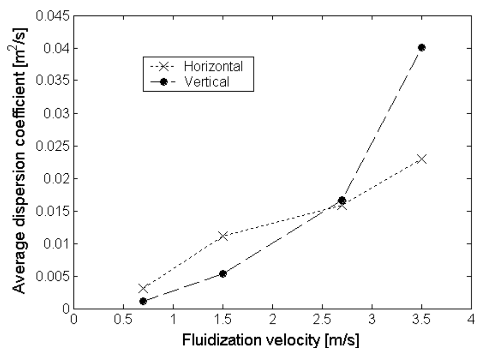

Figure 4: Influence of fluidization velocity on dispersion (Runs A, B, C and D).

The tracer concentration and main velocity fields for Run A are shown in Figure 5, which is characterized by a highly heterogeneous distribution of the tracer concentration field, compared to all other runs. Yet, this regime yields four horizontally aligned vortexes induced by two main bubble paths at approximately the same horizontal locations as in Run C in Figure 3 in spite of the fact that the fluidization velocity is four times as high in Run C. Run A gives high-concentration regions at the downflowing interface of the two central vortexes, specially at the height equal to the bottom bed height, *i.e.* there is a tendency for the tracer particle to remain most of the time at the dense bed surface while moving horizontally (thus leading to a poor vertical dispersion compared to the other runs, as indicated by the dispersion coefficients). In addition, comparison of runs A and C (and B and D, although not shown here) confirms increasing tendencies of the tracer particle to occupy higher locations in the splash zone as fluidization velocity is increased resulting in more uniform distributions of the concentration and velocity in both directions.

In agreement with the trends shown in the previous work (0.40 m wide bed in (12)) and confirmed in an FB boiler (17), the horizontal dispersion measured is enhanced as the solids inventory in the riser is increased (as seen by comparing the results for runs F and C in Table 2). This is explained by the fact that an increased solids inventory in the riser leads to a higher dense bottom bed, which extends a bit further the bubble rising path, ending in faster and more vigorous bubbles reaching the dense bed surface. Even though a higher solids inventory in the riser of an FB boiler helps homogenizing the fuel concentration over the cross section, thus improving oxygen-fuel contact, it also represents a higher operating (fan power) cost.

The two different tracer particle sizes gave no significant difference in horizontal dispersion (see results for runs C-E and F-H), while the vertical dispersion is enhanced when decreasing tracer particle size in the case with the high-ΔP distributor (runs C-E).

Finally, the low-ΔP distributor gave a significantly different mixing pattern compared to the high-ΔP distributor. This can be seen from Figure 6 corresponding to the low-ΔP distributor in Run G and which shows an asymmetry in the tracer concentration field, with high values in the left side and low ones to the right, specially at x=1.05 m (whereas Run F – not shown here - yields an even distribution in tracer concentration over the cross section). The x=1.05 m position corresponds to the location of a high throughflow region, although bed was apparently fluidized over the entire cross-section. In repeated experiments, the location of this

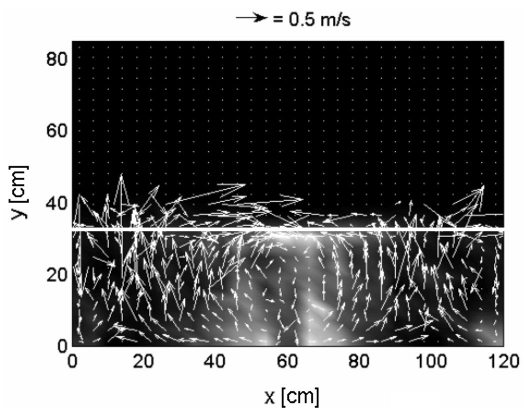

Figure 5: Tracer particle concentration and velocity fields for Run A. White line indicates the bed surface.

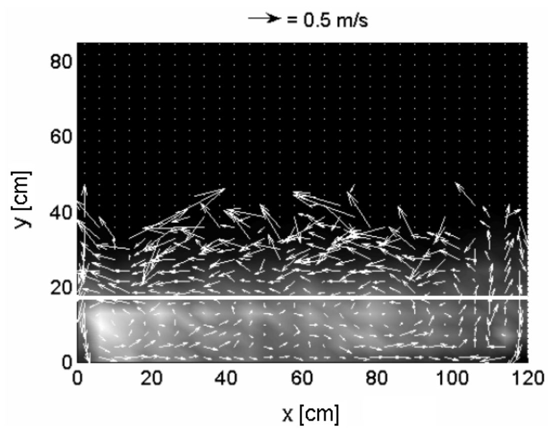

Figure 6: Tracer particle concentration and velocity fields for Run G. White line indicates the bed surface.

throughflow channel varies in a random fashion. The location of this intermittent by-pass strongly affects the flow pattern of the tracer particle: when it is located close to the dense bed surface (y=0.17 to 0.23 m) it will flow away horizontally from the throughflow location and when the tracer particle is located in the bottom of the bed (x=0-5 cm) it will move towards the throughflow location. The large vortex originated by this effect can be seen in Figure 6. In addition, the video recordings show that the tracer moves vertically through the dense bed in locations in the range x=0.10 to 0.80 m, but (as also seen in Figure 6) there are no clear “only-rising” or “only-sinking” regions detected in this range, as is the case with the high-ΔP distributor.

Figure 6 shows a rather uniform tracer concentration field when using a low-ΔP distributor. This would support the use of low-ΔP distributors. However, while this leads to that fuel particles get more evenly distributed, the throughflow effects enhance an uneven distribution of the gas flow (and thereby the oxygen). Thus, despite the high horizontal dispersion values found for tests with the low-ΔP distributor (runs G and I), such conditions may still yield insufficient oxygen-fuel contact.

CONCLUSIONS

A simulated fuel particle inserted in a cold 2D FB unit follows a flow pattern structured into horizontally-aligned vortexes, which is a 2D equivalent of a toroidal structure in 3 dimensions. From a comparison with previous work in a less wide bed than applied in this work it can be concluded that the vortex pattern is induced by the bubble flow and not affected by the bed width.

Increasing the fluidization velocity enhances solids mixing in both vertical and horizontal directions. When the size of the tracer particle is reduced, it reaches both higher velocities within the dense bed and higher locations in the splash zone, increasing vertical dispersion while horizontal dispersion remains more or less the same. An increase in the amount of bed material (constant velocity) enhances the dispersion coefficients by means of a higher dense bottom bed which allows for the formation of larger and faster bubbles. Finally, lowering the air distributor pressure drop significantly changes the mixing pattern, as a result of the formation of high-throughflow (gas) regions which significantly reduces the solids mixing and the gas-fuel contact.

ACKNOWLEDGMENTS

The authors gratefully acknowledge financial support from the Swedish Energy Agency and the EU Research Fund for Coal and Steel under contract RFC-CR-03001.

NOTATION

A	Area [m^2]	r	Mixing rate [m/s]
C	Normalized concentration [m^{-2}]	u_{mf}	Minimum fluidization velocity [m/s]
D	Local dispersion coefficient [m^2/s]	u_t	Particle terminal velocity [m/s]
$\overline{D}$	Averaged dispersion coefficient [m^2/s]	k	Subindex for direction
Da	Damköhler number [-]		
L^*	Characteristic length [m]	Δl	Tracer displacement [m]
n	Subindex for pixel	Δt	Time interval [s]
N	No. of pixels	τ	Characteristic time [s]

REFERENCES

1. Ito, O., Kawabe, R., Miyamoto, T., Orita, H., Mizumoto,M., Miyadera, H., Tomuro, J., Hokari, N., Iwase, T. 1999. *Proc.15th FBC Conf, Savannah,* 217.
2. Highley, J., Merrick, D. 1971. *AIChE Symposium Series* 67, 219.
3. Enwald, H., Almstedt, A.-E., Peirano, E. 1996. *Int. J. Multiphase Flow* 22, 21-66.
4. van Wachem, B. G. M., van der Schaaf, J., Schouten, J. C., Krishna, R., van den Bleek, C. M. 2001. *Powder Technol.,* 116(2-3), 155-165.
5. Peirano, E., Delloume, V., Johnsson, F., Leckner, B., Simonin, O. 2002. *Powder Technol.,* 122(1), 69-82.
6. Ibsen, C.H., Helland, E., Hjertager, B.H., Solberg, T., Tadrist, L., Occelli, R. 2004. *Powder Technol.,* 149 (1), 29-41.
7. Berres, S., Bürger, R., Tory, E.M. 2005. *Applied Math. Modelling* 29, 159-193.
8. Tanskanen, V. 2005. "CFD study of penetration and mixing of fuel in a CFB furnace". MSc Thesis, *Lappeenranta Univ. of Tech. (Finland)*.
9. Niklasson, F., Thunman, H., Johnsson, F., Leckner, B. 2002. *Ind. & Eng. Chem. Res.,* 41(18), 4663-4673.
10. Xiang, Q., Huang, G., Ni, M., Cen, K., Tao, T. 1987. *Proc. 9th FBC*, *Boston*, 546.
11. Schlichthaerle, P., Werther, J. 2001. *Powder Technol.,* 120(1-2), 21-33.
12. Pallarès, D., Johnsson, F. 2006. *Chem. Eng. Sci.,* 61, 2710-2720.
13. Stein, M., Ding, Y. L., Seville, J. P. K., 2002. *Chem. Eng. Sci.,* 57, 3649-3658.
14. Lim, K.S., Agarwal, P.K. 1994. *Chem. Eng. Sci.,* 49, 421-424.
15. Shen, L., Johnsson, F., Leckner, B. 2004. *Chem. Eng. Sci.,* 59, 2607-2617.
16. Rios, G.M., Dang, K., Masson, H. 1986. *Chem. Eng. Comm.,* 47, 247-272.
17. Niklasson, F., Pallarès, D., Johnsson, F. 2006. *Proc. 19th FBC Conf.,* Vienna.