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Modeling Fuel Mixing in a Fluidized Bed
Combustor

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MODELING FUEL MIXING IN A FLUIDIZED BED COMBUSTOR

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ABSTRACT

This paper presents a 3-dimensional model for fuel mixing in fluidized bed combustors. The model accounts for the mixing patterns experimentally shown to govern the mixing in the different zones of the riser and the return leg and can be applied both under bubbling and circulating regimes. Thus, the semi-empirical basis of the model was previously validated in different large-scale fluidized bed combustors and is combined with a model for fuel particle conversion to obtain the fuel concentration. Results obtained with the model are compared with experimental data from the Chalmers 12 MW_{th} CFB combustor, yielding reasonable agreement.

INTRODUCTION

Fuel mixing has a great influence on the overall performance of fluidized bed (FB) combustors. The better the horizontal mixing of the fuel the more homogenous the local stoichiometric ratio over the cross section of the furnace, which should lower the risk of occurrence of locations with un-reacted fuel or oxygen. In the vertical direction a high mixing rate is important in order to secure long enough contact time between the oxygen and the fuel particles. Moreover, good mixing is a prerequisite for an even distribution of heat and gas release from the fuel. Despite the importance of fuel mixing, there is at present a lack of mechanistic models to describe the fuel mixing process and research works have so far been limited either to calculate experimental values of the dispersion coefficient (1, 2), which was only occasionally correlated to operational parameters. Neither has there been much attempt to incorporate mixing of fuel particles in CFD calculations as done in (3). The aim of this paper is to provide a model for fuel mixing in FB combustors to be used in comprehensive fluidized-bed models (4-6) as well as providing basis for future incorporation in CFD modeling.

The continuous physical changes of fuel particles (such as size and density, and thereby terminal velocity) as they undergo conversion (*i.e.* drying, devolatilization and char combustion) lead to constantly changing fuel mixing behaviours. Shortly after the injection into an FB combustor a fuel particle is more likely to occupy the bottom part of the riser due to its relatively large size and high density, while it has an increasing tendency to populate the upper freeboard or even being entrained to the return leg as it gets closer to its burn-out time (due to the smaller size and lower density). Thus, a fuel particle conversion model is required to be combined with the

fuel mixing model in order to describe the transient fluid dynamics resulting from the constantly changing physical properties of a batch of fuel particles (or, if starting out from the fuel conversion modeling, a fluid dynamical model is required to describe the mixing of fuel and oxygen). The present work combines a 3D fuel mixing model with modeling the fuel conversion applying a 3-dimensional meshing for dividing the furnace into cells.

THEORY

Fuel mixing model

In general terms, mixing of fuel particles can be modeled on a semi-empirical basis using the same principles and equations as for the fluid dynamics of the inert (bulk) particles. Literature on macroscopical fluid dynamics of large-scale FB combustors (cf. 4-6) more or less have the same description of the overall solids pattern in the furnace: In the bottom part a dense bottom bed is established which, by means of bubble explosions, form a splash zone of erupted solids in the form of clusters (*cluster phase*) which follow a ballistic movement (back-mixing of solids). Under circulating conditions, a fraction of the solids (*disperse phase*) is entrained higher up in the furnace creating a core-annulus structure in the freeboard, with upflow in the core, net horizontal flow from the core to the downflowing annulus (i.e. most of the back-mixing occurs at the walls). Yet, both the cluster and the disperse phase can exist all the way up through the furnace, with the cluster phase dominating in the bottom freeboard (splash zone) and the disperse phase in the upper freeboard (transport zone). Figure 1 illustrates the main solids movements in the furnace. Table 1 lists the assumptions made with respect to the governing mixing mechanisms in the horizontal and vertical directions for the bottom bed and for the two phases of the freeboard (cluster and disperse phase). These fuel mixing mechanisms are further described below.

The 3-dimensional mesh used to discretize the furnace applies a finer mesh in regions known to exhibit large gradients (e.g. splash zone). At present, perfect mixing is assumed in the return leg and this is therefore not discretized. Note that, at present stage, the fuel mixing model does not account for interactions with the inert solids flow, i.e. fuel mixing is modeled independently from the bulk flow. However, integration of a particle interaction model (such as the one given in (7)) is expected to have a limited influence on the results, since only the fluid dynamics of the smallest fuel particles (which represent a low percentage of the fuel inventory) is expected to be significantly influenced by particle interactions. Future modeling will account for interactions between fuel and bulk particles.

Perfect vertical mixing is assumed in the bottom bed, which is a reasonable assumption according to data in (8) showing that mixing in this region is much faster in the vertical than in the horizontal direction, especially considering that dense bottom beds are usually no higher than a few decimeters compared to several meters in the horizontal direction.

The ballistic mixing of the cluster phase leads to an exponential decay in vertical solids concentration of the fuel phase. Several experimental correlations for this decay constant have been proposed in literature for FB combustors operated under bubbling as well as circulating conditions (9-11). Based on data from several large-scale FB combustors, the cluster phase decay constant was correlated in (12) as:

$$a = 4u_t / u_0 \quad (1)$$

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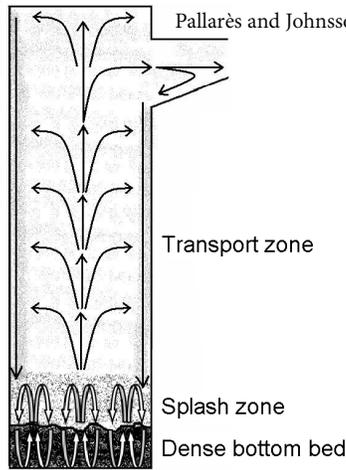


Figure 1: Zones of a CFB riser with corresponding mixing patterns

Although the real mixing mechanism is strongly convective, the experimentally verified existence of a toroidal flow structure around main bubble paths (13, 14) makes it practical to macroscopically simulate the horizontal mixing in the bottom part of the riser in analogy with a diffusion process. This is a common method in literature (see e.g. 15). Thus, the diffusion equation reads:

$$\frac{\partial C}{\partial t} = D \cdot \nabla^2 C + S \quad (2)$$

with S being the source term. Different experimental values and correlations for the horizontal diffusion coefficient D exist in literature (cf. 2) with values of D differing up to two orders of magnitude between investigations carried out under similar conditions. There is little data for beds representative for FB combustors. Two exceptions are given in (2) and (16) where, based on combustor experiments, D is estimated to be about $0.1 \text{ m}^2/\text{s}$ under similar conditions.

The core-annulus flow structure of the dispersed phase (cf. 17, 18) is assumed to follow an exponential decay in solids concentration with height in the furnace in the upflowing core region (solids are assumed to flow at slip velocity), with a decay constant given in (12) as:

$$K = 0.23 / (u_0 - u_t) \quad (3)$$

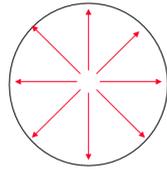
The decreasing fuel upflow with height is due to a net lateral flow from core to wall region (Fig. 2), i.e. fuel back mixing is assumed to mainly occur at the furnace walls. Thus, since the model of this work accounts for heterogeneities in fuel concentration in the core region, wall cells closer to core cells with high fuel concentration get higher incoming fuel flows from the core. The core-to-annulus fuel flow feeds the downflow in the wall region with fuel particles all along the riser. When reaching the bottom bed, this downflow rejoins the dense bed and its diffusive mixing process, which is embodied by the source term S in Eq. (2).

Finally, some of the upflowing particles in the core region which reach the level of the exit duct experience a backflow effect, through which only a certain fraction of these particles reach the cyclone. A proper modeling of this effect requires the use of CFD tools, but experimental correlations for estimating the backflow in standard exit configurations was given in (4, 17-19). From these correlations, an entrainment probability p can be estimated. Since the correlations were found in CFB combustors with a single exit duct an assumption has to be made in order to handle a more general case where n exit ducts exist. Moreover, the backflow effect can be assumed to take place in two steps in series: 1.- some of the upflowing particles in the core follow the gas flow and leave the core upflow towards the exit duct and 2.- once in the exit duct, some particles follow the gas stream all the way into the cyclone, while the rest fall back down into the riser and join the downflow in

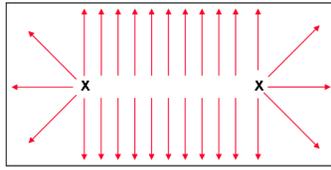
Table 1: Solids mixing mechanisms in an FB furnace applied in this work.

	Mixing mechanism	
	Vertical	Horizontal
Bottom bed	Perfect mixing	Diffusion
Cluster phase	Ballistic	
Disperse phase	Core/annulus	

the wall-layers. These two steps can be expressed as probabilities p_1 and p_2 , as illustrated in Fig. 3.



a) Circular cross section



b) Rectangular cross section

Figure 2: Horizontal core to wall net solids transfer in the freeboard

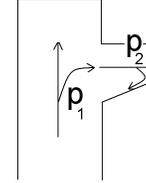


Figure 3: Mechanisms in the backflow effect

Thus, with the single-exit entrainment probability p given by experimental correlations and knowing that $p^* = p_1 \cdot p_2$, the assumption $p_1 = p_2$, leads to the values of p_1 and p_2 given by Eq. (5):

$$p^* = 1 - (1 - p)^n \quad (4) \quad \text{with} \quad p_1 = p_2 = \sqrt{p^*} \quad (5)$$

A detailed description of the arithmetic for solving the solids mixing model in the freeboard using the approach given above can be found in (4).

As far as the externally recirculated fuel particles are concerned, their residence time in the return leg is calculated as the sum of the residence times in the cyclone, downcomer and particle seal. The residence time of fuel particles in the cyclone can be calculated according to (20). The fuel residence time in the downcomer and particle seal is easily calculated with the assumption that it equals the residence time of the bulk solids in the return leg according to Eq. (6).

Several methods to estimate the net solids circulating flow (required in this calculation) in industrial CFB combustors are listed in (21).

$$T_{downcomer\&seal} = V_{fluidized\ inreturnleg} \cdot \rho_s \cdot \left(1 - \varepsilon_{return\ leg}\right) / F_{net,s} \quad (6)$$

Finally, after flowing through the return leg, fuel particles (if not burned out) are refed into the bottom part of the riser. This is implemented through the source term S in Eq. (2).

Fuel conversion model

Several fuel conversion models are available in literature providing results in good agreement with experimental data while being of different approaches and levels of complexity. Since modeling of the fuel conversion is not the focus of this work, attention in this paper is restricted to the outputs required as inputs to the fuel mixing model, namely the loss of mass (Fig. 4) and change in size and density of the fuel particles during conversion. The fuel conversion model presented in (22) is chosen in the present work since it yields low calculation times and yet is shown to give satisfactory agreement with experimental data. The values of the size and density allow calculation at any time step of the terminal velocity (given in normalized values in Fig. 5). Depending on the fuel type, fuel fragmentation might have a large influence on the results and should be taken into account by the fuel conversion model (the assumption on fragmentation applied in this work is given below).

In each time step in the simulation of the conversion of the fuel particle, the data provided by the fuel conversion model in Fig. 4 is used to convert fuel concentration values expressed on a fuel particle basis [particles/m³] to a mass basis [kg/m³]. The progress of the fuel particle terminal velocity shown in Fig. 5 enables the mixing model given in the previous section to account for the continuous change in

the fluid dynamical properties of the fuel particle.
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Method for a steady-state solution

Obviously, modeling the behavior of a batch of fuel particles requires a transient simulation during the whole burnout time. An efficient way to proceed for the case of fuel particles in FB combustors consists in modeling the mixing of a batch of fuel particles during all its burn-out time while ensuring that the amount of fuel in the modeled batch equals the fuel mass that is fed to the FB combustor under continuous fuel feeding during a period equal to the time interval used in the time discretization of the fuel mixing model. This is illustrated in Fig. 6. Applying the above given fuel mixing model, the fuel distribution (in [particles/m³]) in each time step can be calculated using the corresponding value of the terminal velocity. Applying then the pertinent value of the fuel particle mass, the concentration field can be expressed in [kg/m³]. Thus, the spatial distribution of the fuel concentration C originated by the fuel batch at any time step t_i is known (given in the first row in Fig. 6). Having this, a continuous feeding of fuel into an FB combustor can be simulated by a 'continuous batch' approach, in which a new batch is fed to the unit at each time step as illustrated in Fig. 6. In the procedure shown in Fig. 6, the resulting total fuel concentration in each time step is obtained by summing the values in the corresponding column.

As seen from Fig. 6, a stationary concentration field is obtained after a time interval equal to the burnout time of the fuel particles, and this value is equal to the sum of all intermediate values in each time step. However, note that the use of this approach is only applicable to problems where the influence of a certain batch on the other batches can be neglected, which is assumed a reasonable assumption in FB combustors (where fuel represents typically only between 1 and 5% of the total amount of solids in the combustor).

	$t=t_0$	$t=t_1$	$t=t_2$	$t=t_3$...	$t=t_{burnout}$	$t>t_{burnout}$
Batch #1	$C_0(x,y,z)$	$C_1(x,y,z)$	$C_2(x,y,z)$	$C_3(x,y,z)$...	$C_{burnout}(x,y,z)$	0
Batch #2		$C_0(x,y,z)$	$C_1(x,y,z)$	$C_2(x,y,z)$	$C_3(x,y,z)$...	$C_{burnout}(x,y,z)$
Batch #3			$C_0(x,y,z)$	$C_1(x,y,z)$	$C_2(x,y,z)$	$C_3(x,y,z)$...
Batch #4				$C_0(x,y,z)$	$C_1(x,y,z)$	$C_2(x,y,z)$	$C_3(x,y,z)$
...					$C_0(x,y,z)$	$C_1(x,y,z)$	$C_2(x,y,z)$
...						$C_0(x,y,z)$	$C_1(x,y,z)$
...							$C_0(x,y,z)$

Figure 6: The scheme applied to calculate the steady state distribution of fuel concentration.

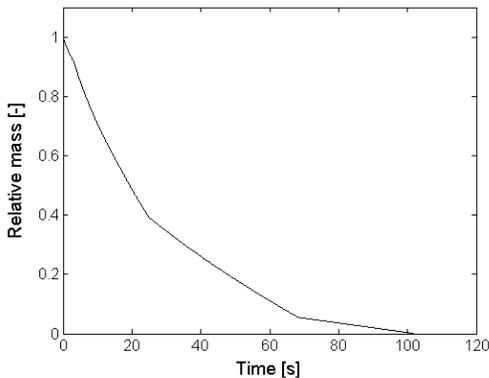


Figure 4: Evolution of the relative mass of a fuel particle

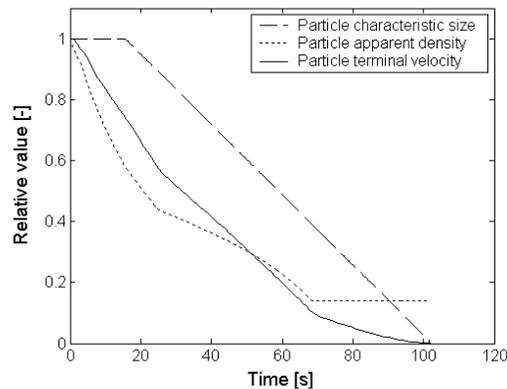


Figure 5: Evolution of the relative size, density and terminal velocity of a fuel particle

RESULTS

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The experimental data used for model validation in this work is from the Chalmers 12 MW_{th} CFB combustor. The size of the combustion chamber is 1.7 m×1.7 m×13.5 m. Two of the walls are covered with refractory lining (0.11 m thick) up to 2 m above the air distributor and the other two are covered all along their height. The fuel is fed by gravity from the fuel chute at a height of 1.1 m above the air distributor (see [23](#) for details). The average height of the dense bottom bed during the tests was estimated (from pressure drop) to be about 0.5 m.

Experimental data on the distribution of fuel concentration under steady-state operation are taken from ([24](#)). Figure 7a shows measured values from 9 points distributed across the cross-section at a low level ($h=0.52$ m) in the furnace. Fuel was Polish coal with a mean size of 8 mm, with moisture, volatiles, char and ash contents (as received) of 17%, 30%, 44% and 9%, respectively. The fuel was fed at a rate of 0.3 kg/s from the fuel chute, placed at the center of the front wall (Fig. 7a), while the inlet of the return leg (fuel re-feed) is located to the right of the rear wall. The data yield fuel concentration values from 23 to 43 kg/m³, with a cross-sectional average of 32 kg/m³. The distribution of the fuel concentration given by the model in the same cross section is shown in Fig. 7b. In the model, fuel particles are assumed to undergo fragmentation after 75% of the devolatilization time, yielding 10 pieces, which is the pattern found in ([25](#)) for coal (although it should be emphasized that fragmentation is strongly fuel-dependent and difficult to model).

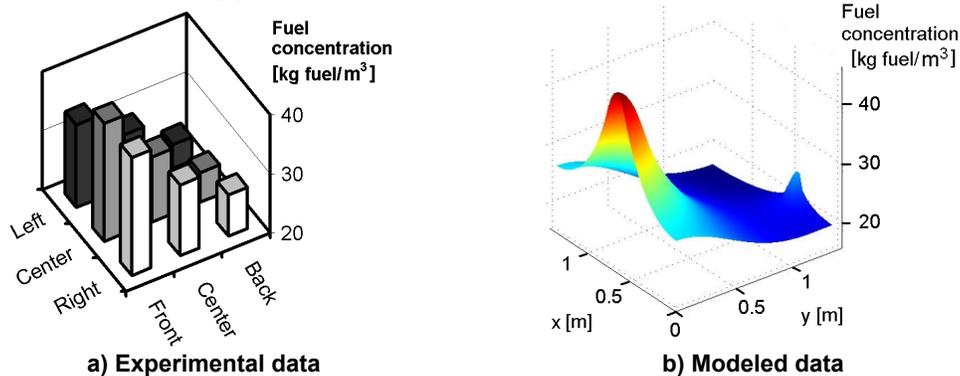


Figure 7: Fuel concentration at bottom region of the Chalmers combustor

In the modeled fuel distribution, fuel concentration values range from 17 to 37 kg/m³, with a cross-sectional averaged value of 25 kg/m³. The average error between modeled and experimental values is 19%.

A comparison between experimental and simulated values at the 9 sampling points is shown in Fig. 8, where a general tendency of the model to slightly underestimate fuel concentration values is observed in all points. There may be several reasons for this (or a combination of them), but the possible overestimation of the fragmentation should be mentioned, since this would lead to an overestimation of the fuel conversion rate and thereby resulting in a lower fuel inventory.

Figure 9 shows the modeled fuel concentration values for the same test at a height of 7 m above the air distributor. As seen, the fuel concentration in the core is much lower than in the walls (in agreement with literature, e.g. [26](#)), where corner effects are significant. It is also seen that the wall with the fuel chute (to the left) gives higher fuel concentration values than at the other walls.

During the first seconds in the furnace a fuel particle releases moisture and stays mostly in the bottom region of the riser. As conversion progresses, volatiles and

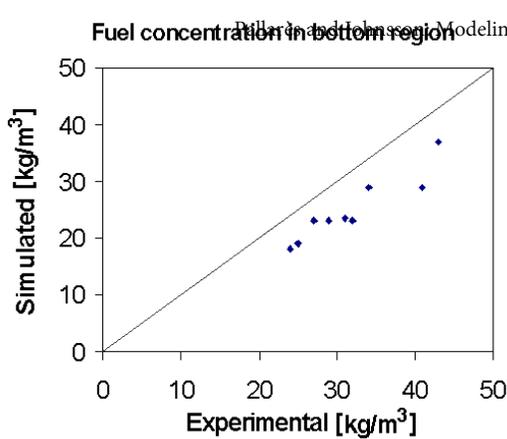


Figure 8: Comparison between modeled and experimental values on fuel concentration

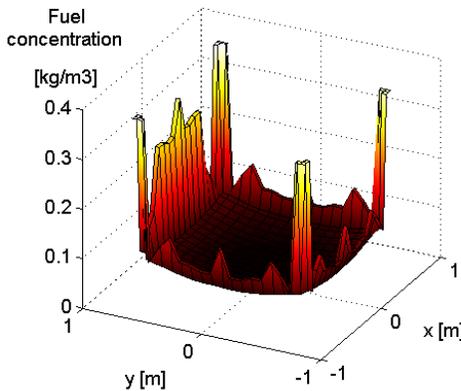


Figure 9: Modeled fuel concentration at a height of 7m

later on char combustion products are released and the fuel particles tend to occupy more of their time at higher locations in the riser.

Simulated cumulative releases of moisture, volatiles and char combustion products are plotted in Fig. 10, showing that, in this example, 95% of the moisture and 79% of the volatiles are expected to be released below a height of 4 m in the riser whereas only 60% of the char combustion products are produced below his level. This reflects the importance of accounting for changes in size and density of fuel particles.

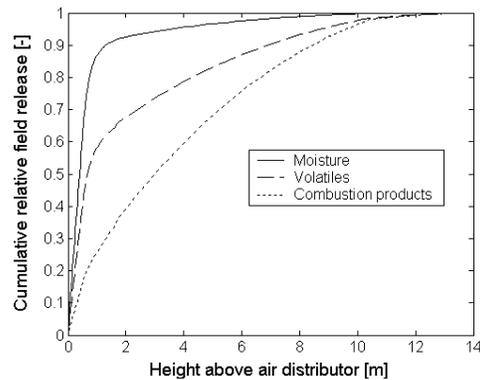


Figure 10: Modeled cumulative fuel field releases

CONCLUSIONS

A model for solids mixing in a fluidized bed has been applied to fuel particles in combination with a fuel conversion model in order to account for physical changes that fuel particles undergo during conversion. The modeled fuel concentrations differ around 20% compared to experimental data. Fuel fragmentation is found to have a large influence on the results and its proper modeling is thereby crucial for a more accurate modeling.

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NOTATION

- | | | | |
|-----|-----------------------------------|----------|--|
| a | Decay constant [m^{-1}] | ρ_1 | Disengagement probability at the riser |
| C | Fuel concentration [kg/m^3] | ρ_2 | Entrainment probability at the duct |
| D | Diffusion coefficient [m^2/s] | T | Residence time [s] |
| F | Solids flow [kg/m^3] | u_0 | Fluidization velocity [m/s] |

K	Decay constant [m^{-1}]	u_t	Terminal velocity [m/s]
n	No. of exit ducts	V	Volume [m^3]
ρ	Entrainment probability with 1 exit duct	ε	Voidage
ρ^*	Entrainment probability with n exit ducts	ρ_s	Solids density [kg/m^3]

REFERENCES

- Schlichthaerle, P., Werther, J. 2001. *Powder Technol.*, 120(1-2), 21-33.
- Niklasson, F., Thunman, H., Johnsson, F., Leckner, B. 2002. *Ind. & Eng. Chem. Res.*, 41(18), 4663-4673.
- Tanskanen, V. 2005. "CFD study of penetration and mixing of fuel in a CFB furnace". Master Thesis, *Dept. of Energy and Environment, Lappeenranta Univ. of Tech.* (Finland).
- Pallarès, D., Johnsson, F. 2006. *Progress in Energy and Combustion Science*, 32 (5), 539-569.
- Lei, H. 1998. "Hydrodynamic modelling of circulating fluidized bed reactors", PhD Thesis. *Tokyo Univ. of Agric. and Tech.* (Japan).
- Hannes, J. 1996. "Mathematical Modelling of Circulating Fluidized Bed Combustion". *City-Print Verlag GmbH. Aachen* (Germany).
- Palchonok, G., Breitholtz, C., Thunman, H., Leckner, B. 1997. "Impact of heat and mass transfer on combustion of a fuel particle in CFB boilers". *Proc. 14th FBC Conf.*, 871-888.
- Bellgardt, D., Werther, J. 1986. *Powder Technol.*, 48 (2), 173-180.
- Lewis, W.K., Gilliland, E.R., Lang, P.M. 1962. *Chem. Eng. Prog. Symp. Ser.*, 58, 65-78.
- Kunii, D., Levenspiel, O. 1990. *Powder Technol.* 61(2), 193-206.
- Lei, H., Horio, M. 1998. *J. of Chem. Eng. of Japan* 31(1), 83-94.
- Johnsson F, Leckner B, 1995. *Proc. 13th Int. Conf. on FBC*, 671-679.
- Pallarès, D., Johnsson, F. 2006. *Chem. Eng. Sci.*, 61, 2710-2720.
- Xiao, P., Yan, G., Wang, D. 1998. *J. of Thermal Science* 7 (2), 78.
- Berruti, F., Scott, D.S., Rhodes, E. *Can. J. of Chem. Eng.*, 64 (1), 48-56.
- Xiang, Q., Huang, G., Ni, M., Cen, K., Tao, T. 1987. *Proc. of 9th Int. Conf. on FBC*, 546.
- Werther J. 1993. *Circulating Fluidized Bed Technology IV*, 1-14.
- Sanderson, W.E. 1994. "The separation effect on the CFBC's riser exit". *Delft University Press Science*, TU-Delft (The Netherlands).
- Werdermann, C.C. 1992. "Feststoffbewegung und Wärmeübergang in zirkulierenden Wirbelschichten von Kohlekraftwerken".
- Kang, S.K., Kwon, T.W., Kim, S.D. 1998. *Powder Technology* 58 (3), 211-220
- Edvardsson, E., Åmand, L.-E., Thunman, H., Leckner, B., Johnsson, F. 2006. *Proc. of the 19th Int. Conf. on FBC*, Vienna.
- Palchonok, G., Didalenko, V., Stanchits, L., Borodulya, V., Werther, J., Leckner, B. 1997. *Proc 14th FBC Conf., Vancouver*.
- Leckner, B., Golriz, M.R., Zhang, W., Andersson, B.Å., Johnsson, F. 1991. *Proc. 11th FBC Conf*, 771-776.
- Åmand, L.E., Lyngfelt, A., Karlsson, M., Leckner, B., 1997. Internal report no. A 7-221. *Dept. of Energy Conversion, Chalmers Univ. of Tech.* (Sweden).
- Zhang, H., Cen, K., Yan, J., Ni, M. 2002. *Fuel*, 81 (14), 1835-1840 .
- Zhang, W., Johnsson, F., Leckner, B. 1995. *Chem. Eng. Sci.*, 50(2), 201-210. ⁸