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The impact of bed particle size in heat transfer to membrane walls of supercritical CFB boiler

ARTUR BŁASZCZUK[1*]
WOJCIECH NOWAK[1]
SZYMON JAGODZIK[2]

[1] Czestochowa University of Technology, Institute of Advanced Energy Technologies, Dąbrowskiego 73, 42-200 Czestochowa, Poland

[2] Tauron Generation S.A., Lagisza Power Plant, Pokoju 14, 42-250 Bedzin, Poland

Abstract Experimental research has been carried out in a supercritical circulating fluidized bed combustor in order to indicate the effect of the bed particle size on bed-to-wall heat transfer coefficient. The bed inventory used were 0.219, 0.246 and 0.411 mm Sauter mean particles diameter . The operating parameters of a circulating fluidized bed combustor covered a range from 3.13 to 5.11 m/s for superficial gas velocity, 23.7 to 26.2 kg/(m^{2}s) for the circulation rate of solids, 0.33 for the secondary air fraction and 7500 to 8440 Pa pressure drop. Furthermore, the bed temperature, suspension density and the main parameters of cluster renewal approach were treated as experimental variables along the furnace height. The cluster renewal approach was used in order to predict the bed-to-wall heat transfer coefficient. A simple semi-empirical method was proposed to estimate the overall heat transfer coefficient inside the furnace as a function of particle size and suspension density. The computationally obtained results were compared with the experimental data of this work.

Keywords: Heat transfer coefficient; Cluster renewal approach; Bed particle size; Suspension density; Circulating fluidized bed

*Corresponding Author. E-mail: ablaszczuk@is.pcz.czest.pl

Nomenclature

c	–	specific heat, kJ/(kg K)
d	–	diameter, m
d_p	–	mean bed particle size, mm
D_h	–	hydraulic diameter, m
e	–	emissivity
f	–	fractional of the wall covered by clusters
g	–	gravitational acceleration, m/s^2
G_s	–	solids circulation flux, kg/(m^{2}s)
h	–	bed-to-wall heat transfer coefficient, W/(m^{2}K)
H	–	height of furnace, m
k	–	thermal conductivity, W/(m K)
L_c	–	cluster characteristic travel length, m
n	–	empirical constant
p	–	pressure, Pa
Pr	–	Prandtl number, $\mathrm{Pr} = \frac{c_g \mu_g}{k_g}$
SA	–	secondary air fraction
t_c	–	cluster residence time, s
T	–	temperature, K, °C
U_c	–	cluster descent velocity, m/s
U_{mf}	–	minimum fluidization velocity, m/s
U_o	–	superficial gas velocity, m/s
U_t	–	terminal velocity of bed solid particles, m/s
z	–	vertical coordinate

Greek symbols

ε	–	voidage
ε_{mf}	–	voidage at minimum fluidization
δ	–	nondimensional gas layer thickness between the wall and cluster
ρ	–	density, kg/m^3
ρ_s	–	suspension density, kg/m^3
σ	–	Stefan-Bolzmann's constant, W/(m^{2}K^4)

Subscripts

av	–	average
b	–	bed
c	–	cluster
cal	–	calculated value
con	–	convection component of heat transfer coefficient
d	–	dispersed
exp	–	experimental value
f	–	fluidization

g	–	gas
m	–	minimum
max	–	maximum
p	–	particle
rad	–	radiation component of heat transfer coefficient
rc	–	radiation component of heat transfer coefficient from cluster phase
rd	–	radiation component of heat transfer coefficient from dishpersed phase
ref	–	reference value
s	–	suspension
sat	–	saturation
w	–	wall

1 Introduction

In the last twenty years the construction of circulating fluidized bed (CFB) combustors has been developed with progressively increasing capacity of the CFB units. The use of circulating fluidized bed technology with regard to heat and power generation is the most common one due to such factors as fuel flexibility, high efficiency, compact furnace size, good heat transfer characteristics, efficient combustion, adaptability to load change and also low emission of pollutants. Some detailed information on the current status of circulating fluidized bed technology in different regions of the world was presented [1,2]. In literature [3–5], there are some mathematical models of heat transfer and flow phenomena in the furnace chamber of boilers. Heat transfer data from CFB units in a large-scale are rarely presented due to commercial reasons and measuring difficulties. Until now, only a limited number of studies [6–11] have treated the effect of bed particle size on the heat transfer coefficient at the wall surface in a laboratory scale at low furnace temperatures ranging from 65 °C to 400 °C. Moreover, several experimental works on heat transfer inside the furnace chamber [12–18] were carried out in circulating fluidized beds with different mean particle diameters, at bulk temperature in the measurement section ranging from 700 °C to 900 °C. These results obtained by different authors [6–18] confirm that larger particles give a lower overall bed-to-wall heat transfer coefficient. A proper size of bed particles inside the combustion chamber makes it possible to obtain the vertical distribution of suspension density necessary to maintain the optimum operating temperature which improves its performance.

In this study, the effect of bed particle size on the heat transfer coefficient between a submerged vertical membrane wall and a fluidized bed was determined. The bed inventory used in the present study were solid particles with Sauter mean diameter of 0.219, 0.246 and 0.411 mm. A semiempirical heat transfer model was proposed to predict the overall heat transfer coefficient at different sizes of bed particles. The cluster renewal approach was used to estimate a bed-to-wall heat transfer coefficient inside the furnace chamber. The predicted values of that heat transfer coefficient were compared with experimental data for CFB unit in a large-scale.

2 Cluster renewal model of heat transfer in CFB furnace

In the present model the hydrodynamic parameters of circulating fluidized bed system have been taken into account. In CFB boilers the gas-solid flow pattern above the refractory line (i.e., the secondary air injection level) inside the furnace chamber is typically of a core-annulus structure. In the core with a dilute suspension, solids move upward with an occasional presence of clusters. The gas velocity in the dilute core is well above the superficial gas velocity, while that in the annulus is low to negative since solids in the form of clusters move downwards. The downward velocity of clusters in the annulus region is between 2 and 8 m/s in large-scale CFB boilers [19]. In a certain distance from the membrane wall, outside of the thin gas layer, the particles form clusters and then solid agglomerates disintegrate as shown in Fig. 1.

The gas gap thickness, δ, between the wall and the cluster is calculated using an expression given by Lints *et al.* [20]. Based on these considerations, the membrane wall comes interchangeably in contact with clusters and the dispersed phase. A typical cluster travel length, L_c, and residence time, t_c, (i.e., time of contact between the cluster and the wall) depend upon bed hydrodynamic conditions and they can be estimated in a CFB unit by the expressions given by Wu *et al.* [21]

$$L_c = 0.0178\rho_b^{0.596} \ , \tag{1}$$

and proposed by authors [22]

$$t_c = \frac{L_c}{U_c} = \frac{L_c}{0.75\left(\rho_p g d_p/\rho_g\right)^{0.5}} \ , \tag{2}$$

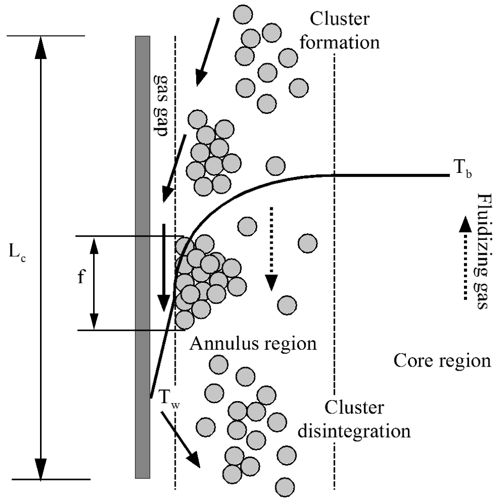

Figure 1: Conceptual view of cluster and gas gap close to a membrane-wall [20].

where U_c represents cluster descent velocity, ρ_b is bed density, ρ_p particle density, ρ_g gas density, d_p denotes mean bed particle diameter in m and g is gravitational acceleration.

In applied mechanistic model, the heat transfer coefficient between the membrane wall and the bed includes contributions of particle convection, h_p, gas convection, h_g, cluster convection, h_{con}, gas conduction, h_w and radiation h_{rad} are considered to be relevant. The radiation heat transfer coefficient is a combination of radiation from the clusters h_{rc} and from the dispersed phase h_{rd}. Thus, in circulating fluidized bed combustors the overall bed-to-wall heat transfer coefficient above the secondary air injection level in dilute phase is given by the expression

$$h = h_{con} + h_{rad} = f h_p + (1 - f)\, h_g + f h_{rc} + (1 - f)\, h_{rd} \,, \qquad (3)$$

where f denotes the fractional wall coverage by clusters, Duta *et al.* [23]:

$$f = 1 - \exp\left[-4300\,(1 - \varepsilon)^{1.39}\,(D_h/H)^{0.22}\right] \,. \qquad (4)$$

In the Eq. (6), ε denotes voidage, D_h is a hydraulic diameter and H represents the furnace height in m. The derivation of the coefficients in (6) was estimated on basis of data from several commercial CFB boilers. The heat transfer Eqs. (7)–(11) used in this mechanistic model are given below and the formulas needed for estimating the physical and thermal properties of the cluster as well as a typical flue gas of a coal-fired fluidized bed can be found in [24].

The particle convective heat transfer coefficient, h_p, is given by expression

$$h_p = \frac{1}{\left(\frac{1}{h_c} + \frac{1}{h_w}\right)} = \frac{1}{\left(\frac{\pi t_c}{4 k_c \rho_c c_c}\right)^{0.5} + \frac{d_p \delta}{k_g}} , \qquad (5)$$

where h_c denotes the heat transfer coefficient from the clusters to the membrane wall, h_w means the heat transfer through the gas layer between cluster and the wall due to conduction, k_c represents cluster thermal conductivity, ρ_c denotes cluster density, c_c is the specific heat of clusters, $d_p\delta$ means gas layer thickness between the wall and the cluster, and k_g is gas thermal conductivity. The gas convection heat transfer coefficient is given as

$$h_g = \frac{k_g c_p}{d_p c_g} \left(\frac{\rho_d}{\rho_p}\right)^{0.3} \left(\frac{U_t^2}{g d_p}\right)^{0.21} \mathrm{Pr} , \qquad (6)$$

where k_g is thermal conductivity of gas, U_t represents terminal velocity of the bed particles, ρ_d denotes dispersed phase density, Pr is Prandtl number, c_g and c_c are the specific heat for gas and cluster, respectively. Similar to the convective heat transfer coefficient, the radiation from the bed to the wall comprised two components: (i) a cluster phase h_{rc} and (ii) a dispersed phase h_{rd}. The cluster phase component is

$$h_{rc} = \frac{\sigma \left(T_c^2 + T_w^2\right) \left(T_c^+ T_w\right)}{\left(1/e_c + 1/e_w - 1\right)} , \qquad (7)$$

where T_c represents the temperature of cluster, T_w denotes the wall temperature and σ is the Stefan-Bolzmann's constant. In the Eq. (9), wall emissivity e_w equal to 0.8 and particle emissivity $e_p = 0.7$, as suggested by Basu [24]. With a particle emissivity of 0.7, the emissivity of cluster e_c is estimated from the formulae, which takes account of the inert particle radiation:

$$e_c = 0.5\left(1 + e_p\right) . \qquad (8)$$

Radiation between a dispersed phase and the wall is estimated from the expression

$$h_{rd} = \frac{\sigma \left(T_b^2 + T_w^2\right) \left(T_b + T_w\right)}{\left(1/e_d + 1/e_w - 1\right)} , \qquad (9)$$

where T_b represents bed temperature, e_d is emissivity of that dispersed phase. For commercial CFB boilers, emissivity of the dispersed phase is estimated by correlation

$$e_d = \left[\frac{e_p}{(1-e_p)\,0.5}\left(\frac{e_p}{(1-e_p)\,0.5}+2\right)\right]^{0.5} - \frac{e_p}{(1-e_p)\,0.5}\ . \qquad (10)$$

3 Semiempirical heat transfer equations

In the present work, a few empirical relationships between the overall bed-to-wall heat transfer coefficient, suspension density and particle size are adapted to estimate the heat transfer coefficient (HTC) for the given Sauter mean particle diameter from a reference HTC. A numerical quantity of heat transfer coefficient is calculated for another particle diameter and both local suspension density in the CFB furnace and a size correction factor as the following are taken into consideration [9]:

$$h = h_{ref}\left(\frac{d_{ref}}{d_p}\right)^n\ , \qquad (11)$$

where the subscript *ref* indicates the reference data and n means the empirical constant. The reference particle diameter d_{ref} equals 0.179 mm, as suggested by [9] and the corresponding bed-to-wall heat transfer coefficient is estimated from the following equation proposed by the authors of this work:

$$h_{ref} = 3.38 + 49.16\rho_b - 0.30\rho_b^2\ , \qquad (12)$$

where ρ_b refers to the bed density in kg/m^3. The above empirical relationship was obtained on the basis of furnace data (about 100 data points) for a supercritical circulating fluidized bed boiler.

The parameter n is a function of suspension density and plays a more significant role than the mean particle size suggested in [9]. In that case the exponent in the empirical correlation (11) is calculated from the following equation:

$$n = 2.218\times10^{-1} - 1.069\times10^{-3}\rho_s + 2.338\times10^{-4}\rho_s^2 - 1.788\times10^{-6}\rho_s^3\ , \qquad (13)$$

where ρ_s is the suspension density. The expression (13) is valid to bed particles of diameter in the range from 0.179 mm to 0.545 mm. In this heat transfer study, the coefficient n was obtained for the same data which was used for the comparison in the Section 5.

4 Performance test and experimental conditions

Experiments were carried out under a steady state condition on a commercial circulating fluidized bed combustor in order to analyse the effect of the bed particle size on the bed-to-wall heat transfer coefficient under similar operating conditions. The main process data of the CFB facility are summarized in Tab. 1

Table 1: Experimental conditions referred to in this heat transfer study.

Parameter	Unit of measure	Overall range
Thermal capacity, Q_{th}	MW_{th}	384–966
Superficial gas velocity, U_o	m/s	3.13–5.11
Terminal velocity, U_t	m/s	1.99–2.91
Minimum fluidization velocity, U_{mf}	m/s	0.01643–0.05443
Particle density, ρ_p	kg/m^3	2680–2750
Bed pressure, p_b	Pa	7050–8000
Pressure drop, Δp	Pa	7500–8440
Bed temperature, T_b	K	1118–1201
Furnace temperature difference, ΔT	K	18–55

The heat transfer coefficient was estimated on the basis of performance tests on 1296 t/h supercritical CFB combustor at Tauron Generation S.A. Lagisza Power Plant in Poland. These performance tests were conducted in a supercritical CFB boiler with a bed cross section of 27.6 m×5.3 m and the furnace being 48 m tall. The membrane walls consist of tubes connected by the fins. The tubes have a 51 mm outer diameter and a 52 mm pitch (Fig. 2). These active heat transfer surfaces were placed above the refractory line in CFB furnace. Additional details of CFB facility are reported elsewhere [25,26].

To measure the bed temperature K-type thermocouples were used. They were installed on the front wall of the combustion chamber. Four set of thermocouples were placed along the vertical direction in this furnace chamber with equal spacing at the nondimensional distance z/H of 0.25, 0.50, 0.65 and 0.87 (Fig. 2). Moreover, in the present study the wall temperature of the heat transfer surface was estimated, as suggested by Basu [24] by

$$T_w = T_{sat} + 30 , \tag{14}$$

where T_{sat} denotes the saturation temperature of steam. At the CFB unit, the steam temperature inside the membrane wall was continuously measured and stored as 5-minute averages using the online monitoring system.

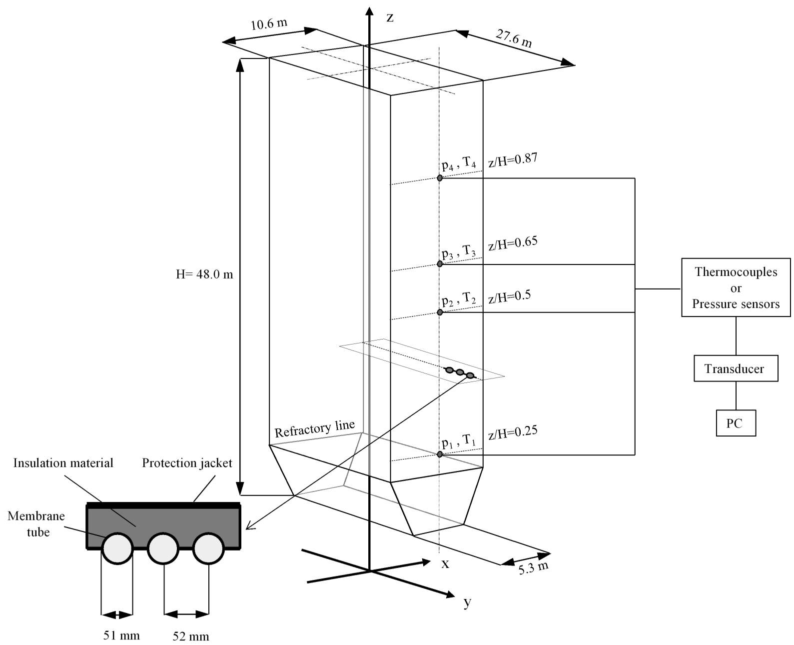

Figure 2: The furnace chamber of CFB boiler and locations of multiple sampling ports with schematic diagram of measurement system.

The saturation temperature of the fluid inside the tube is slightly lower than the wall surface temperature.

In a supercritical CFB boiler, pressure measurements were carried out at the walls of the combustion chamber. Static pressure measurements were carried out at the same non-dimensional distances z/H from the air distributor like in the case of the furnace temperature measurements (Fig. 2). Pressure taps (4 pieces at the front wall) were mounted flush with the wall of furnace chamber to measure pressure drops around the water membrane wall and then connected to differential pressure transducer. From the measured pressure drop along CFB furnace, the apparent cross-sectional bed average voidage, ε, was estimated as

$$\varepsilon = 1 - \frac{1}{g\rho_p}\left(\frac{\Delta p}{\Delta H}\right) , \tag{15}$$

where Δp represents pressure drop, ΔH denotes distance between pressure tappings.

All performance tests were conducted with five sizes of bed inventory, with Sauter mean particle diameters of 0.219, 0.246 and 0.411 mm. The mean bed particle size was defined as the bed material size for the bottom ash since the ports for representative bed material sampling along the height were not installed in CFB furnace. The particle size distributions (PSD) of the bed are provided from the performance test and the sieve analysis (Fig. 3).

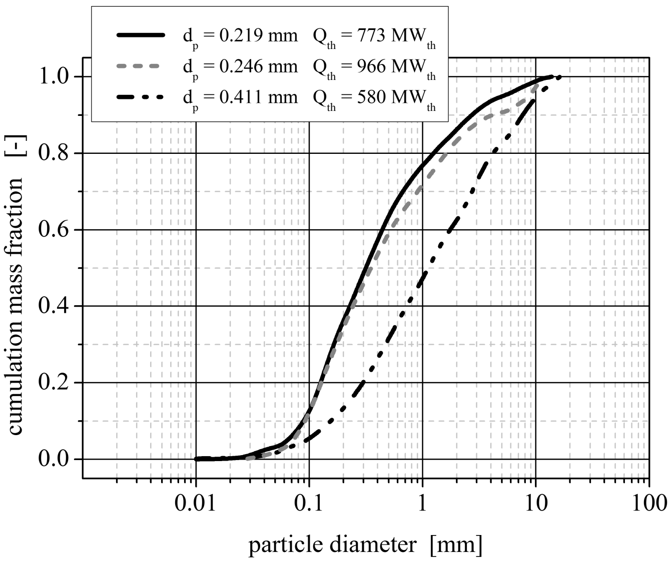

Figure 3: Particle size distributions of bed inventory during performance tests.

5 Results and discussion

Figure 4 demonstrates the effects of mean bed particle size on the bed-to-wall heat transfer coefficient for different relative furnace heights of CFB boiler. It can be observed that the present study confirms strong dependency of the bed-to-wall heat transfer coefficient on the bed particle sizes and the suspension density. With an increase in the furnace height, the solid suspension density decreased and as a result the values of heat transfer coefficient decreased. As presented in Fig. 4, some differences in the overall heat transfer coefficient between the particle size of 0.219 mm and 0.411 mm were at the level of +98% on average. One can notice that CFB heat transfer is more significant for smaller particles and large suspension density. This effect was also confirmed by the authors [9]. Moreover, variations of the overall heat transfer coefficient as a function of the bed particle sizes have an exponential function. The fitting curves corresponding to all relative heights of furnace are drawn in Fig. 4. The heat transfer coefficient is predicted by Eq. (5) with the correlation coefficient covering the range from 0.91 to 0.98. The reported values of heat transfer coefficient in a function of particle diameter are a satisfactory correlation for relative heights of furnace 0.25, 0.65 and 0.87. The correlation of heat transfer coefficient for $z/H = 0.5$ is not satisfactory. Obtained results of the heat transfer coefficient by using the cluster renewal approach are characterized by low uncertainties. The uncertainty in the heat transfer coefficient is in the range from ± 7.9 to ± 21.8 W/(m^{2}K) and for the mean bed particle size is from ± 0.00004 m to ± 0.00008 m. The major source of inherent error is the uncertainty in the heat transfer coefficient caused by the level of accuracy of the sieve analysis of bed inventory, as shown in Fig. 4, the uncertainty in a mean bed particle diameter is caused by attrition process of bed particles during the sieve analysis of each bed material sample. Smaller bed particle sizes have a larger contact area with the membrane wall when compared with bigger particles. Operating finer bed particles could be compared to a denser curtain. This curtain is located between the high temperature core and the wall. It contributes to a decrease in the net radiation flux.

The bed particle size has an impact on the heat transfer process, especially on the suspension radiation absorption and scattering. Finally, more heat is transported from finer particles to the gas and then conducted from the gas to the wall. Smaller bed particles within the mixture of gas-solids make the gas gap thinner and hence resulting in lower gas gap conduction resistance. Consequently, the particle heat transfer mechanism plays more important role in the case of smaller size bed particles rather than the bigger ones. In a large-scale CFB unit, finer bed particles contribute to higher overall bed-to-wall heat transfer coefficient values due to an increase in the convective component of the heat transfer process. As far as two convective heat transfer components are concerned, with a decrease in the bed particle size, the particle convection heat transfer coefficient increased rapidly while the gas convection heat transfer was almost unchanged.

It is commonly known that the bed-to-wall heat transfer coefficient is strongly correlated with the suspension density inside CFB furnace. Figure 5 presents the bed-to-wall heat transfer coefficient as a function of rela-

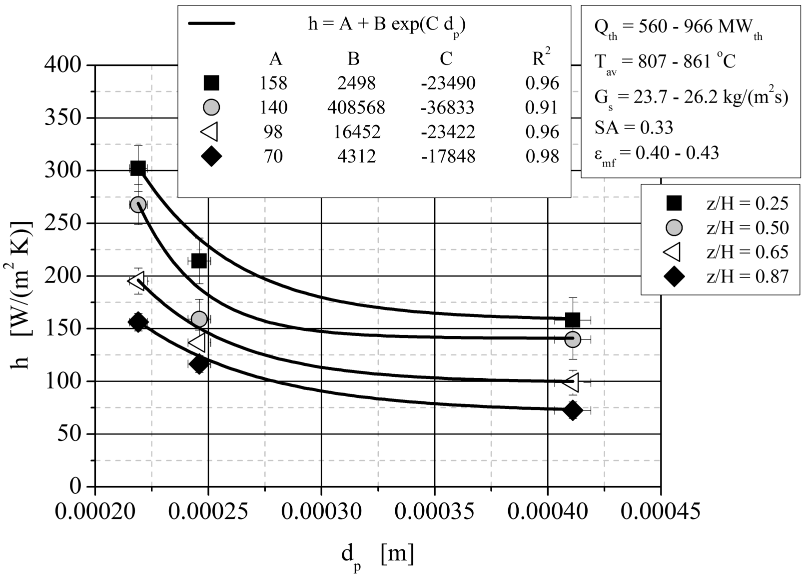

Figure 4: Variation of local heat transfer coefficient along the CFB furnace height versus mean bed particle size.

tive suspension density for different mean bed particles sizes. In this work, relative suspension density results are given in a dimensionless scale and referred to the maximum value of the suspension density inside the combustion chamber during all performance tests. The maximum value of solids suspension density equals 6.22 kg/m^3 during performance tests. The overall heat transfer coefficient increases with suspension density as observed in a supercritical circulating fluidized bed boiler in a large-scale. At suspension density less than 2.93 kg/m^3 the heat transfer from the dispersed phase to the wall is dominated by thermal radiation, as distinct from the region of CFB furnace with high solid concentration. Then, particle heat transfer coefficient plays a major role in the heat transfer mechanism under dense phase conditions whereas the wall is covered by more clusters (about f = 57%). In a dilute region the membrane wall is covered by only a few clusters (about f = 12%). That was quite predictable since thermal capacity of the bed inventory is much higher for higher values of suspension density.

The heat transfer data in Fig. 5 indicate an essential influence of suspension density on the heat transfer coefficient. The slope of experimental data is proportional to suspension density and also to the bed tempera-

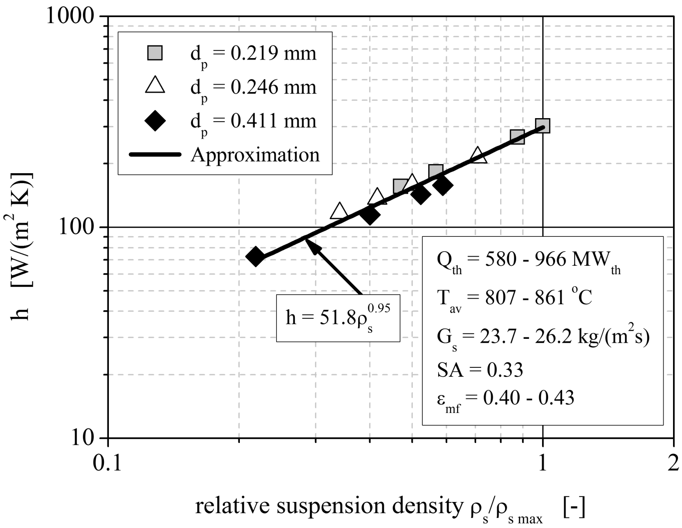

Figure 5: Effect of suspension density on bed-to-wall heat transfer coefficient for different mean bed particle size.

ture. That tendency of variation was proportional to the fractional power of suspension density and it depended upon the local fluid characteristic and the thermal boundary layer. A thin gas gap reduced the bed-to-wall heat transfer coefficient, especially the particle heat transfer coefficient. It could be explained by the fact that the gas gap provides a significant amount of resistance to the cluster heat transfer, even if it is quite small in thickness. Otherwise, during performance tests the thickness of the gas layer varied along the height of the combustion chamber from 0.227 mm at dense phase conditions (bottom region) to 1.205 mm at dilute phase conditions (exit region). With a decrease in suspension density, the cluster heat transfer coefficient decreases, however, the gas layer thickness increases. High suspension density results in resistance of the low thickness gas layer between clusters and an active heat transfer surface and also in an increase in the cluster heat transfer coefficient. This in turn leads to a higher bed-to-wall heat transfer, which has been confirmed in [27,28].

The accuracy of the proposed Eq. (12) is proved via comparing computationally obtained results with the results of experimental tests. In this comparison, the same input variables (i.e., mean bed particle size and suspension density) are used in the experimental tests as the calculations.

Figure 6 shows the comparison between the overall bed-to-wall heat transfer coefficient calculated from Eq. (13) and estimated on the basis of the performance tests at different values of mean bed particle sizes: 0.219, 0.246, and 0.411 m. The whole comparison covers the numerical value normally occurring in circulating fluidized bed systems. The middle line at 45° is the line of perfect agreement and the remaining two dashed lines show the boundary of 21%. Typically, the bed-to-wall heat transfer coefficient decreased when the bed particle size and the CFB furnace height increased.

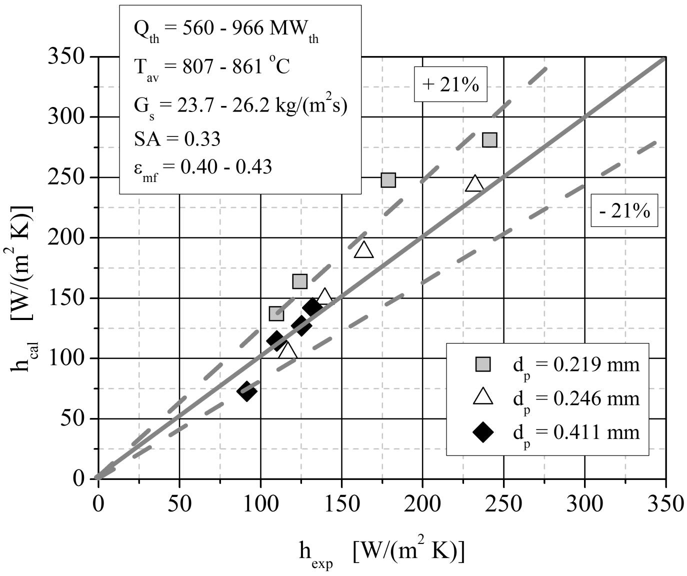

Figure 6: Comparison of predicted overall bed-to-wall heat transfer coefficient with experimental data for different mean bed particle size.

At bed particle size of 0.219 mm inside the combustion chamber with suspension density more or equal to 5.44 kg/m^3, the variations of heat transfer data from the theoretical line were high as opposed to the results obtained for other sizes of bed particles. This suggests that the heat transfer coefficient increases continuously together with an increase in the operating pressure. The same tendency of heat transfer coefficient variations was reported by others [29]. It is observed in Fig. 6 that good correspondence with the theoretical prediction was obtained in the case of two bed particle sizes in the present study, 0.246 mm and 0.411 mm. Deviation between computationally obtained results and experimental data exceed 21%. It indicates that as far as the proposed heat transfer correlation is concerned, the prognosis is promising and it is possible to estimate the bed-to-wall heat transfer coefficient reliably. Moreover, determined empirical correlation confirms both flexibility and sufficient accuracy of the heat transfer model.

6 Summary

The experimental investigation on the heat transfer behavior inside a furnace chamber for the large-scale CFB facility shows the following key results:

- The overall heat transfer coefficient is strongly dependent on three important factors: (i) bed particle size, (ii) suspension density, (iii) and hydrodynamics of a circulating fluidized bed.
- In the bed, finer particles produce higher heat transfer coefficient than bigger ones due to a higher contact point on the active heat transfer surface.
- The experimental heat transfer data indicate that the bed-to-wall heat transfer coefficient is reduced along the height of a supercritical CFB unit furnace.
- Suspension density plays an essential role in the heat transfer behavior since particle convection is a dominant mechanism of the heat transfer between the bed inventory and the membrane wall surface.
- The bed-to-wall heat transfer coefficient generally increased with the suspension density due to high heat capacity of solids.
- A semiempirical method is proposed to predict the overall heat transfer coefficient in a circulating fluidized bed. The proposed method gives a satisfactory description regarding the effect of the size of bed particles on the overall heat transfer coefficient. The heat transfer Eq. (12) can be successfully applied to fluidized beds of particle with Sauter mean diameters ranging from 0.219 mm to 0.411 mm. The obtained 21% discrepancy between the experimental and calculated values of the overall heat transfer coefficient seems satisfactory.

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References

[1] Hupa M.: *Current status and challenges within fluidized bed combustion.* Advanced Combustion and Aero-thermal Technologies NATO Sciences for Peace and Security Series C: Environ. Secu. **1**(2007), 87–101.

[2] Koornneef J., JungingerM., Faaij A.: *Development of fluidized bed combustion – An overview of trends, performance and cost.* Prog. Energ. Combust. Sci. **33**(2007), 1, 19–55.

[3] Zima W., Grądziel S., Cebula A.: *Modeling of heat and flow phenomena occurring in water wall tubes of boilers for supercritical steam parameters.* Arch. Thermodyn. **31**(2010), 3, 19–36.

[4] Taler J., Taler D., Sobota T., Dzierwa P.: *New technique of the local heat flux measurement in combustion chambers of steam boilers.* Arch. Thermodyn. **32**(2011), 3, 103–116.

[5] Taler J., Taler D., Kowal A.: *Measurements of absorbed heat flux and water-side heat transfer coefficient in water wall tubes.* Arch. Thermodyn. **32**(2011), 1, 77–88.

[6] Andersson B.: *Effects of bed particle size on heat transfer in circulating fluidized bed boilers.* Powder Technol. **87**(1996), 3, 239–248.

[7] Bis Z.: *CFB boilers. Theory and practice.* Czestochowa University Press, Czestochowa 2010 (in Polish).

[8] Basu P., Nag P.K.: *An investigation into heat transfer in circulating fluidized beds.* Int. J. Heat Mass Tran. **30**(1987), 11, 2399–2409.

[9] Pagliuso J.S., Lombardi G., Goldstein L.: *Experiments on the local heat transfer characteristics of a circulating fluidized bed.* Exp. Therm. Fluid Sci. **20**(2000), 3-4 , 170–179.

[10] Gungor A.: *A study on the effects of operational parameters on bed-to-wall heat transfer.* Appl. Therm. Eng. **29**(2009), 11-12, 2280–2288.

[11] Xie D., Bowen B.D., Grace J.R., Lim C.J.: *Two-dimensional model of heat transfer in circulating fluidized beds. Part II: Heat transfer in a high density CFB and sensitivity analysis.* Int. J. Heat Mass Tran. **46**(2003), 12, 2193–2205.

[12] Werdermann C.C., Werther J.: *Heat transfer in large-scale circulating fluidized bed combustors of different sizes.* In: Circulating Fluidized Bed Technology IV (A.A. Avidan Ed.), AIChE, New York 1994, 428–435.

[13] Golriz M.R., Sunden B.: *An analytical-empirical model to predict heat transfer coefficient in circulating fluidized bed combustors.* Heat Mass Transfer **30**(1995), 377–383.

[14] Wu R.L., Grace J.R., Lim C.J., Brereton C.: *Suspension-to-surface heat transfer in a circulating fluidized bed combustor.* AIChE J., **35**(1989), 10, 1685–1691.

[15] LUAN W., LIM C.J., BRERETON C.M.H., BOWEN B.D., GRACE J.R.: *Experimental and theoretical study of total and radiative heat transfer in circulating fluidized beds*. Chem. Eng. Sci. **54**(1999), 17, 3749–3764.

[16] HAN G.Y., CHO Y.J.: *Radiative heat transfer in a circulating fluidized bed coal combustor*. Powder Technol. **102**(1999), 3, 266–273.

[17] ERIKSSON M., GORLITZ M.R.: *Radiation heat transfer in circulating fluidized bed combustors*. Int. J. Therm. Sci. **44**(2005), 4, 399–409.

[18] SHI D., NICOLAI R., REH L.: *Wall-to-bed heat transfer in circulating fluidized bed boilers*. Chem. Eng. Process. **37**(1998), 287–293.

[19] WERTHER J.: *Fluid dynamics, temperature and concentration fields in large-scale CFB combustors*. In: Circulating Fluidized Bed Technology – VIII. International Academic Publishers, Beijing 2005, 1–25.

[20] NIRMAL VIJAY G., REDDY B.V.: *Effect of dilute and dense phase operating conditions on bed-to-wall heat transfer mechanism in a circulating fluidized bed combustor*. Int. J. Heat Mass Tran. **48**(2005), 3275–3283.

[21] WU R.L., GRACE J.R., LIM C.J.: *A model for heat transfer in circulating fluidized beds*. Chem. Eng. Sci. **45**(1990), 12, 3389–3398.

[22] NOYMER P.D., GLICKSMAN L.R.: *Descent velocities of particle clusters at the wall of a circulating fluidized bed*. Chem. Eng. Sci. **55**(2000), 22, 5283–5289.

[23] DUTTA A., BASU P.: *An improved cluster renewal model for estimation of heat transfer coefficient on the water-walls of commercial circulating fluidized bed boilers*. J. Heat Trans. – T. ASME **126**(2004), 6, 1040–1043.

[24] BASU P.: *Combustion and gasification in fluidized beds*. Taylor & Francis Group, 2006.

[25] BLASZCZUK A., LESZCZYNSKI J., NOWAK W.: *Simulation model of the mass balance in a supercritical circulating fluidized bed combustor*. Powder Technol. **246**(2013), 313–326.

[26] BLASZCZUK A., NOWAK W., JAGODZIK SZ.: *Effects of operating conditions on deNO$_x$ system efficiency in supercritical circulating fluidized bed boiler*. J. Power Technol. **93**(2013), 1, 1–8.

[27] REDDY B.V., BASU P.: *A model for heat transfer in a pressurized circulating fluidized bed furnace*. Int. J. Heat Mass Tran. **44**(2001), 15, 2877–2887.

[28] REDDY B.V.: *Fundamental heat transfer mechanism between bed-to-membrane water-walls in circulating fluidized bed combustors*. Int. J. Energy Res. **27**(2003), 9, 813–824.

[29] GUPTA A.V.S.S.K.S., NAG P.K.: *Bed-to-wall heat transfer behavior in a pressurized circulating fluidized bed*. Int. J. Heat Mass Tran. **45**(2002), 16, 3429–3436.