PERGAMON

International Journal of Heat and Mass Transfer 42 (1999) 4065–4075

International Journal of HEAT and MASS TRANSFER

www.elsevier.com/locate/ijhmt

Scaling in a circulating fluidized bed: particle concentration and heat transfer coefficient in a transport zone

Yu.S. Teplitskiy[a,*], G.A. Ryabov[b]

[a]A.V. Luikov Heat and Mass Transfer Institute, National Academy of Sciences of Belarus, Minsk 220072, Republic of Belarus
[b]All-Russia Thermal Engineering Institute, Moscow 109280, Russia

Received 17 January 1997

Abstract

Similarity analysis of heat and mass transfer processes in circulating fluidized beds (CFB) which operate according to Catalytic Cracking CFB and CFB Combustor schemes is performed. On this basis the rules of transport processes scaling are formulated in the form of a system of dimensionless governing parameters. Particle concentration distribution and heat transfer coefficient in a transport zone of the riser with a height 6.6 m are studied experimentally. The obtained and literature data on the values of ρ and α_{c-c} are generalized on the basis of the developed system of dimensionless groups.

1. Introduction

Problems of scaling, which in practice are understood as determination of a minimum quantity of dimensionless groups composed of dimensional independent variables and fully determining similarity of both transport processes in a CFB and of CFB proper, are very important and attract the attention of researchers. On the basis of a cluster model of a flow at a riser wall, Horio et al. [1] obtained four groups of parameters which in a dimensionless form are [2]:

$$Fr_D,\quad \frac{\rho_s}{\rho_f},\quad \frac{u}{u_t},\quad \frac{J_s}{\rho_s u}. \tag{1}$$

Neglecting an inertia term in the Ergun equation, the authors of [3] determined a system of four groups which resembles (1):

$$Fr_D,\quad \frac{\rho_s}{\rho_f},\quad \frac{u}{u_{mf}},\quad \frac{J_s}{\rho_s u}. \tag{2}$$

Taking into account the presence of u_{mf} in (2), one should prefer system (1) which is more physical in nature for a CFB.

Nondimensionalizing governing differential equations for gas and particles the authors of [4] obtained system of five dimensionless groups:

$$Fr_D,\quad \frac{\rho_s}{\rho_f},\quad \frac{d}{D},\quad Re_d,\quad \frac{J_s}{\rho_s u}. \tag{3}$$

Account for balance between viscous and inertia forces in (3) by introducing Re_d makes the system rather universal. Chang and Louge [5] found another system of five dimensionless groups:

$$Fr^*\quad \frac{\rho_s}{\rho_f},\quad \frac{D}{d\theta},\quad Ar^*\quad \frac{J_s}{\rho_s u}, \tag{4}$$

which allows also for nonsphericity of particles. A rather limited experience of use of systems (3) and (4)

* Corresponding author.

0017-9310/99/$ - see front matter
PII: S0017-9310(99)00048-4

Nomenclature

a_f	thermal diffusivity of gas
$Ar = gd^3(\rho_s/\rho_f - 1)/\nu_f^2$	Archimedes number; $Ar^* = Ar\theta^3$
c	specific heat capacity
d	particle diameter
$Fo = a_f t/d^2$	Fourier number
$Fr_D = u^2/gD$, $Fr^* = u^2/gd\theta$	Froude numbers
$Fr_t = (u - u_t)^2/gH$	modernized Froude number
g	gravity acceleration
$Ga = gd^3/\nu_f^2$	Galileo number
h	height above a gas distributor
H	riser height
H_{mf}	bed height at $u = u_{mf}$
$\bar{J}_s = J_s/\rho_s(u - u_t)$	dimensionless solid mass flux
J_s	net specific solid mass flux (mass circulating flux)
L	length of a heat transfer surface
M	solid mass in a riser
$Nu = \alpha d/\lambda_f$	Nusselt number
ΔP	pressure drop
$Pr = c_f \rho_f \nu_f/\lambda_f$	Prandtl number
$Re_d = ud/\nu_f$, $Re_D = uD/\nu_f$	Reynolds numbers
R	residue on a sieve
t	particle residence time at a heat transfer surface
T	temperature
u, u_{mf}, u_t	superficial gas velocity, minimum fluidization velocity, particle terminal velocity
v	downward particle velocity at a riser wall

Greek symbols

α	heat transfer coefficient
ϵ_{mf}	minimum fluidization on bed porosity
θ	coefficient of particle non-sphericity
λ_f	thermal conductivity of gas
ν_f	kinematic viscosity of gas
μ_f	dynamic viscosity of gas
ρ	particle concentration (mean over the horizontal section of a riser)
ρ_w	particle concentration at a riser wall

Subscripts

c–c	conductive–convective
f	gas
mf	minimum fluidization
s	particles
w	heat transfer surface

does not make it possible at present to give preference to any of the systems.

One substantial fact which refers to the applicability of systems (3) and (4) should be mentioned. As is known, one of the main differences of a CFB from a classical bubbling fluidized bed is that an outer circulating particles flux, generated by a system itself when $u > u_t$, exists in a CFB along with developed inner circulation. With respect to a outer circulating particle flux all CFB can be divided into two classes:

1. forced circulating fluidized beds in which J_s is a governing (independent) parameter assigned during CFB operation (Catalytic Cracking CFB (CCCFB));
2. free circulating fluidized beds in which J_s is a determinable (dependent) parameter, i.e. the function of independent parameters (u, d, ρ_s, etc.) (CFB Combustor (CFBC)).

As a rule, these two types of a CFB differ also in the value of the ratio H/D (of about $20 \div 80$ for CCCFB and $5 \div 10$ for CFBC) [6]. Account for different functional determination of J_s in CFB of the types mentioned is absent in systems (3) and (4), thus imposing certain limitations on their practical use.

The main purpose of the present paper is to formulate rather a substantiated and universal complete system of dimensionless groups which reflect similarity of transfer processes in a transport zone of a CFB and allow for the mentioned difference of CFB operating according to Catalytic Cracking CFB and CFB Combustor schemes.

2. Similarity of transfer processes in CFB

Similarity analysis will be made separately for CFB of different types.

2.1. Catalytic Cracking CFB

For an arbitrary local characteristic (FCC) of a CFB of this type the following relation can be written:

$$\mathrm{FCC} = f(g, D, d, H, L, h, J_s, \rho_s, \rho_f, \lambda_f, \rho_f c_f, u, v_f). \tag{5}$$

On the basis of the π-theorem of the dimensional theory a dimensionless analogue of (5) is written:

$$(\mathrm{FCC})' = f\left(\frac{H}{d}, \frac{L}{d}, \frac{h}{d}, Ga, \frac{J_s}{\rho_s(u-u_t)}, \frac{\rho_s}{\rho_f}, Re_d, Re_D, Pr\right). \tag{6}$$

Since in the paper main attention is paid to modeling of processes in a transport zone of a CFB, it is more correct to introduce a dimensionless flux of particles in the form of $J_s/\rho_s(u-u_t)$ rather than $J_s/\rho_s u$. A general criterial Eq. (6) is rather complex and it should be, whenever possible, simplified to a maximum. It is admissible to neglect number Pr because in gas fluidization it changes slightly. A more customary number Ar is used instead of Ga, and, finally, form, and remain only one simplex h/H instead of two simplexes H/d and h/d with the simplex H/d being truncated as having no physical sense. In this case it is supposed that the presence of two independent one-dimensional spaces: 'local', associated with a linear size of the probe (L), and 'global' having a characteristic size equal to the height of the whole riser H. Within the framework of these conceptions, h, which characterizes the probe location in the riser, is the coordinate of 'local' space in 'global' one, in which actually 'local' space is considered as non-dimensional. It is clear that in such a formulation only simplexes h/H and d/L are physically meaningful. As a result of these simplifications, (6) is reduced to the equation

$$(\mathrm{FCC})' = f\left(\frac{h}{H}, \frac{d}{L}, Ar, \bar{J}_s, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}\right). \tag{7}$$

Re_d and Re_D are replaced by equivalent Re_d and d/D for the convenience of comparing (3) and (4).

On the basis of (7) it can be concluded: full similarity of two CFB operating by the CCCFB scheme is realized when five dimensionless groups are equal:

$$Ar, \bar{J}_s, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}. \tag{8}$$

Five groups of (8) give five independent equations for determining 5 dimensional parameters of a laboratory CFB which models a real apparatus (d, D, J_s, u, ρ_s). Full similarity of local transfer processes is realized when seven dimensionless groups are equal:

$$\frac{h}{H}, \frac{d}{L}, Ar, \bar{J}_s, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}. \tag{9}$$

2.2. CFB Combustor

Formally Eq. (5) can also be written for this case. But now, as has been already mentioned, the value of mass circulating flux (J_s) is a function of determining parameters of the system

$$\frac{J_s}{\rho_s(u-u_t)} = f(u-u_t, H, H_{mf}, D, g). \tag{10}$$

Using the theory of vector dimensions in the Huntley form [7] it can easily be shown that (10) is equivalent to the following relation between dimensionless groups

$$\bar{J}_s = A Fr_t^{a_0}\left(\frac{H_{mf}}{H}\right)^{b_0} \tag{11}$$

(the quantity D disappears due to its vector dimension—length along a horizontal, for nondimensionalization of which there is no uniform quantity). Combining (7) and (11), an arbitrary dimensionless characteristic (FCF) of the bed of this type is obtained:

$$(\mathrm{FCF})' = f\left(\frac{h}{H}, \frac{d}{L}, Ar, Fr_t, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}, \frac{H_{mf}}{H}\right). \tag{12}$$

In the case when a certain pressure drop (Δp) is maintained in the riser during operation, the quantity $\Delta P/\rho_s(1-\epsilon_{mf})Hg$ should be used instead of H_{mf}/H.

On the basis of (12) it can be concluded: two CFBs operating according to the CFBC scheme will be similar if the following groups

$$Ar, Fr_t, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}, \frac{H_{mf}}{H} \tag{13}$$

in them are equal to each other, respectively. Six dimensionless groups of (13) give six independent equations for determining six dimensional characteristics of a laboratory CFB (d, D, u, ρ_s, H, H_{mf}). Eight dimensionless groups

$$\frac{h}{H}, \frac{d}{L}, Ar, Fr_t, Re_d, \frac{d}{D}, \frac{\rho_s}{\rho_f}, \frac{H_{mf}}{H} \tag{14}$$

determine similarity of local transfer processes.

Dimensionless groups $\bar{J}_s = J_s/\rho_s(u-u_t)$ and $Fr_t = (u-u_t)^2/gH$ used above were first introduced in [8]. The number $\bar{J}_s$ is a generalized characteristic of catalytic cracking CFB, and Fr_t is a generalized characteristic of CFB Combustor. These numbers have a simple physical meaning: a dimensionless circulating flux $\bar{J}_s$ characterizes particle concentration at the riser

Table 1
Similarity of CFB operating according to the CCCFB scheme

Quantity	CFB1	CFB2	CFB3	CFB4
J_s, kg/m^2 s	50.0	83.2	50.0	15.8
D, m	1.0	0.24	1.0	1.55
u, m/s	6.0	2.57	6.0	7.6
u_t, m/s	1.06	0.45	0.81	1.03
d, m	0.2×10^{-3}	0.048×10^{-3}	0.2×10^{-3}	0.31×10^{-3}
μ_f, kg/m s	449×10^{-7}	179×10^{-7}	365×10^{-7}	179×10^{-7}
ρ_f, kg/m^3	0.31	1.2	4.81	1.2
ρ_s, kg/m^3	2600	10,064	2600	650
	Similarity		Similarity	

exit, and the Froude number Fr_t—the ratio of kinetic energy of particles to their potential energy in the gravity field. A fruitful character of employment of these numbers will be shown in what follows on the examples of the development of the methods for calculating vertical distributions of ρ and α_{c-c} in the CFB riser.

2.3. Calculation of the parameters of laboratory CFB modelling typical installations

Table 1 presents calculated parameters CFB2 and CFB4 of laboratory CFBs ($T = 20°C$; $P = 0.1$ MPa, fluidizing agent—air), which model high-temperature CFBs operating according to the CCCFB scheme: CFB1 ($T = 800°C$; $P = 0.1$ MPa) and CFB3 ($T = 1000°C$; $P = 2$ MPa).

Table 2 gives an analogous calculation for CFBs operating by the CFBC scheme.

2.4. Distribution of particle concentration over the height of a transport zone of a CFB

On the basis of relations (7) and (12) dimensionless dependencies of particle concentration on determining factors can rather easily be obtained.

2.4.1. Catalytic Cracking CFB

A general expression for ρ/ρ_s follows from (7) if it is assumed the function f to be power

$$\frac{\rho}{\rho_s} = B\left(\frac{h}{H}\right)^{a} Ar^{b} \bar{J}_s^{c} Re_d^{d} \left(\frac{d}{D}\right)^{e} \left(\frac{\rho_s}{\rho_f}\right)^{f}. \tag{15}$$

It can be easily shown that (15) is substantially simplified. In fact, ρ can be presented as

$$\rho = \rho(H)(h/H)^{a}. \tag{16}$$

An obvious relation between J_s and particle concentration at the riser exit (ρ_{exit}) is used:

$$J_s = \rho_{exit}(u - u_t). \tag{17}$$

Assuming $\rho(H) = C\rho_{exit}$, the relation for determining ρ is obtained

$$\frac{\rho}{\rho_s} = C\bar{J}_s\left(\frac{h}{H}\right)^{a}, \tag{18}$$

which is a partial case of (15).

Table 2
Similarity of CFB operating according to the CFBC scheme

Quantity	CFB1	CFB2	CFB3	CFB4
H, m	12.0	2.2	12.0	19.2
H_{mf}, m	0.5	0.092	0.5	0.8
D, m	1.0	0.24	1.0	1.55
u, m/s	6.0	2.57	6.0	7.6
u_t, m/s	1.06	0.45	0.81	1.03
d, m	0.2×10^{-3}	0.048×10^{-3}	0.2×10^{-3}	0.31×10^{-3}
μ_f, kg/m s	449×10^{-7}	179×10^{-7}	365×10^{-7}	179×10^{-7}
ρ_f, kg/m^3	0.31	1.2	4.81	1.2
ρ_s, kg/m^3	2600	10,064	2600	650
	Similarity		Similarity	

2.4.2. CFB Combustor

For this case an analogue of (15) follows from (12):

$$\frac{\rho}{\rho_s} = B_1\left(\frac{h}{H}\right)^{a_1} Ar^{b_1} Fr_t^{c_1} Re_d^{d_1}\left(\frac{d}{D}\right)^{e_1}\left(\frac{\rho_s}{\rho_f}\right)^{f_1}\left(\frac{H_{mf}}{H}\right)^{g}. \tag{19}$$

A simplified version of (19) follows from the combination of (11) and (18):

$$\frac{\rho}{\rho_s} = C_1 Fr_t^{a_0}\left(\frac{H_{mf}}{H}\right)^{b_0}\left(\frac{h}{H}\right)^{a_1}. \tag{20}$$

2.5. Conductive–convective heat exchange with a riser wall in a transport zone

In this section the above-obtained Eqs. (7) and (12) are also used.

2.5.1. Catalytic Cracking CFB

On the basis of (7), assuming as earlier the unknown relation to be power-law and one-term, the following is obtained:

$$Nu_{c\text{-}c} = N\left(\frac{h}{H}\right)^{a'} Ar^{b'} J_s^{c'} Re_d^{d'}\left(\frac{d}{D}\right)^{e'}\left(\frac{\rho_s}{\rho_f}\right)^{f'}\left(\frac{d}{L}\right)^{h}. \tag{21}$$

2.5.2. CFB Combustor

Similarly, from (12)

$$Nu_{c\text{-}c} = N_1\left(\frac{h}{H}\right)^{a'_1} Ar^{b'_1} Fr_t^{c'_1} Re_d^{d'_1}\left(\frac{d}{D}\right)^{e'_1}\left(\frac{\rho_s}{\rho_f}\right)^{f'_1} \times\left(\frac{H_{mf}}{H}\right)^{g'}\left(\frac{d}{L}\right)^{h_1}. \tag{22}$$

The obtained dimensionless Eqs. (21) and (22) determine a structural dependence of the coefficient $\alpha_{c\text{-}c}$ on determining parameters of a CFB and will be used in the generalization of experimental data.

3. Experimental study

3.1. Experimental setup

Studies were conducted on a cold model of a boiler with a CFB which is a vertical Plexiglas riser of height 6.6 m and cross-section 0.4 × 0.4 m provided by a cyclon and a system for returning solid particles (Fig. 1). A circulating flow of particles was controlled by varying amount of air supplied to the gate. A fluidizing agent is air at ambient temperature. Sand ($\rho_s = 2470$ kg/m^3) and anionite ($\rho_s = 1240$ kg/m^3) were used as disperse material. Fractional composition of disperse materials is shown in Fig. 2. Experiments were conducted at air velocities ranging from 3.0 to 6.0 m/s. With an initial loading of disperse material 100–150 kg the variation of bed mass in the riser was attained by changing the level of a dense bed in a recirculation loop. The value of circulating particle flux was measured by two different methods: (a) by the velocity of single particle motion in a dense bed at a wall of the recirculation loop; and (b) by cutoff of disperse material in the recirculation loop with the gate during a certain period of time (~10 s) and subsequent weighing. Thus, it was possible to measure bed mass in the riser and the value of a circulating particle flux during the experiments. Concentration of particles was determined by a drop of static pressure in the riser. The points of static pressure sampling are shown in Fig. 1. The coefficient of heat transfer between a circulating fluidized bed and the riser wall was measured by the heat transfer probe

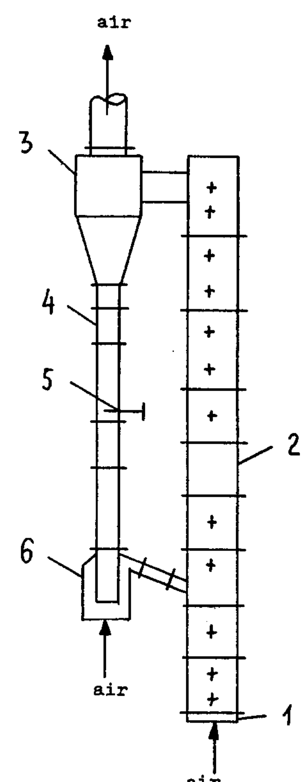

Fig. 1. Experimental setup: 1, gas distributor; 2, riser 0.4 × 0.4 m; 3, cyclon; 4, recirculation column 0.15 × 0.15 m; 5, gate; 6, hydraulic gate; +, pickup of static pressure.

(Fig. 3). The levels of probe location were 3.57 and 5.14 m from the gas distributor. The heat transfer surface of the probe (membrane) was made of foiled fiber-glass with thickness 0.2 mm. It is designed for removing a certain amount of heat, for mounting the probe in the setup and thermal insulation of the probe. A 55-mm-dia disk is a working part of the membrane. In a 2-mm gap between the working part and the remaining portion of the membrane the foil is etched on two sides. The function of the gap is to provide minimum heat losses from the probe to a setup wall. The body of the probe is made of aluminium in the form of a truncated cone with a larger base having a diameter of 55 mm. Its function is distribution and supply of heat from the heater to the heat transfer surface. The wide base of the probe body is glued to the working part of the membrane with a thin layer of epoxy adhesive. Two chromel–alumel thermocouples for measuring temperature are embedded in the probe body. A silicon transistor fixed to the probe body is used as a heater. An analysis of the operation mode of the probe and losses in it indicates that an error of the determination of the heat transfer coefficient does not exceed 10%. The probe was calibrated under the conditions of natural convection.

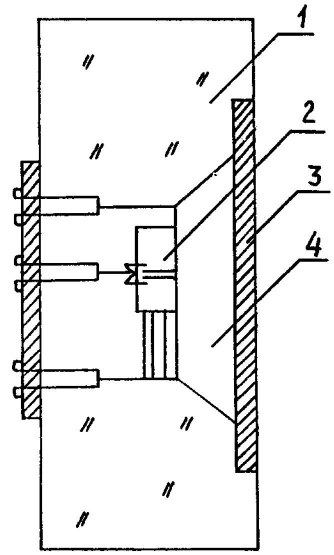

Fig. 3. Heat transfer probe: 1, fragment of a riser wall; 2, KT 8102 transistor; 3, membrane; 4, cone.

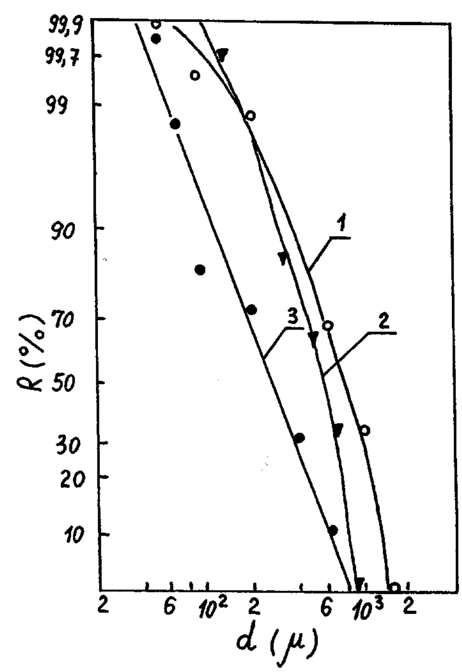

Fig. 2. Fractional composition of disperse materials: 1, anionite $d = 0.650$ mm; 2, anionite $d = 0.550$ mm; 3, sand $d = 0.250$ mm.

3.2. Experimental results

3.2.1. Distribution of particle concentration over the riser height

Figs. 4–6 present typical relations $\rho = \rho(h)$. An air velocity, distance from the gas distributor and mass of the material in the riser substantially affect a value of ρ.

3.2.2. Conductive–convective heat transfer

Fig. 7 presents experimental data on the dependence of the coefficient $\alpha_{c\text{–}c}$ on a mean concentration of particles in a horizontal section of the riser. A substantial growth of $\alpha_{c\text{–}c}$ with an increase in ρ is evident. The effect of particle size on the value of $\alpha_{c\text{–}c}$ is also seen: the coefficient $\alpha_{c\text{–}c}$ decreases with an increase in a mean diameter of particles.

4. Generalization of experimental data

4.1. Distribution of particle concentration over the riser height in a transport zone of a CFB

4.1.1. Catalytic Cracking CFB

The available literature data [9–18] was generalized in accordance with the theoretical relation (18). As a result, its validity was confirmed and a simple equation was obtained for calculating ρ:

$$\left(\frac{\rho}{\rho_s}\right) = \bar{J}_s \left(\frac{h}{H}\right)^{-0.82}. \tag{23}$$

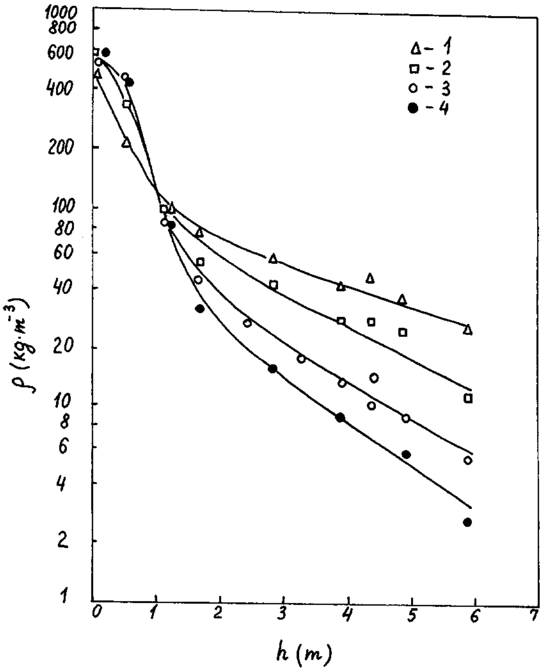

Fig. 4. Influence of air velocity on the profile of particle concentration over the riser height: 1, u = 6.3 m/s; 2, 5.3; 3, 4.3; 4, 3.75. Anionite d = 0.550 mm; M = 70–80 kg.

This equation describes experimental points with a root-mean-square error of 16% (Fig. 8). The range of verification of (23) is: $0.046 \le d \le 0.286$ mm; $5.9 \le J_s \le 147$ kg/m^2 s; $2.7 \le H \le 12$ m. It follows from comparison of (18) and (23) that $C = 1$, thus indicating the equality $\rho(H) = \rho_{\text{exit}}$. On this basis a conclusion can be made about the absence of downward particle flows in the riser zone adjacent to the exit.

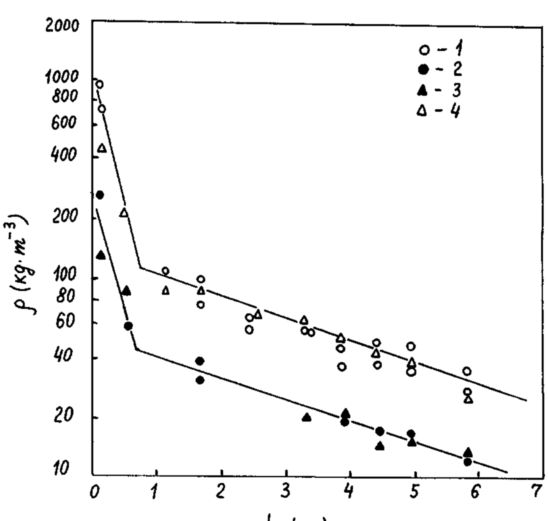

Fig. 5. Influence of bed mass in the riser on particle concentration: 1, M = 70 kg; 2, 35; 3, 20; 4, 8. Anionite d = 0.650 mm; u = 6.7–6.8 m/s.

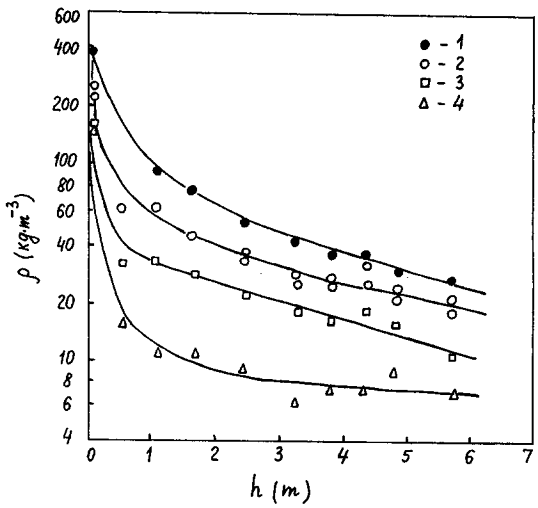

Fig. 6. Variation of particle concentration of sand over the riser height: u = 5.6 m/s; 1, M = 80 kg; J_s=45 kg/m^2 s; 2, 30; 22; 3, 10; 20; 4, 70; 37.

4.1.2. CFB Combustor

The data obtained were generalized by relation (20). As a result the following equation was found

$$\frac{\rho}{\rho_s} = 0.26 Fr_t^{0.55} \left(\frac{H_{\text{mf}}}{H}\right)^{0.80} \left(\frac{h}{H}\right)^{-0.82}, \tag{24}$$

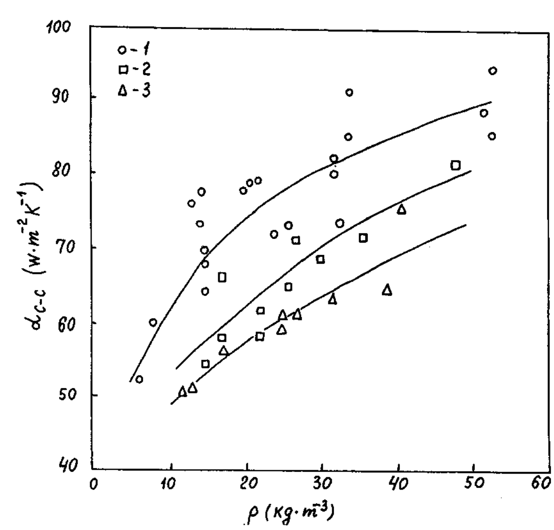

Fig. 7. Dependence of the coefficient of heat transfer on particle concentration: 1, sand d = 0.250 mm; 2, anionite d = 0.550 mm; 3, anionite d = 0.650 mm.

shown in Fig. 9a. Since in the experiments a value of a circulating mass of particles flux was also measured, it turned possible for the conditions of the experiments to obtain the dependence of the type of (23)

$$\frac{\rho}{\rho_s} = 3.0\bar{J}_s\left(\frac{h}{H}\right)^{-0.82}, \tag{25}$$

shown in Fig. 9b. It follows from (25) that $\rho(H) = 3.0\rho_{exit}$, thus indicating the presence of a developed downward flow of particles in the zone $h \approx H$ and reflecting hydrodynamic conditions at the rise exit.

It should be noted that the dependence of ρ on the local height h has no universal character. The data of [19] obtained on an operating boiler with a CFB having a power of 12 MW at constant pressure drop over the riser height was generalized:

$$\frac{\rho}{\rho_s} = 0.053Fr_t^{0.62}\left(\frac{h}{H}\right)^{-0.45}. \tag{26}$$

As is seen, the dependence on h in this case is smoother.

4.2. Conductive–convective heat exchange between a CFB and riser walls in a transport zone

4.2.1. Catalytic Cracking CFB

Experimental data of [9,11,12] were generalized

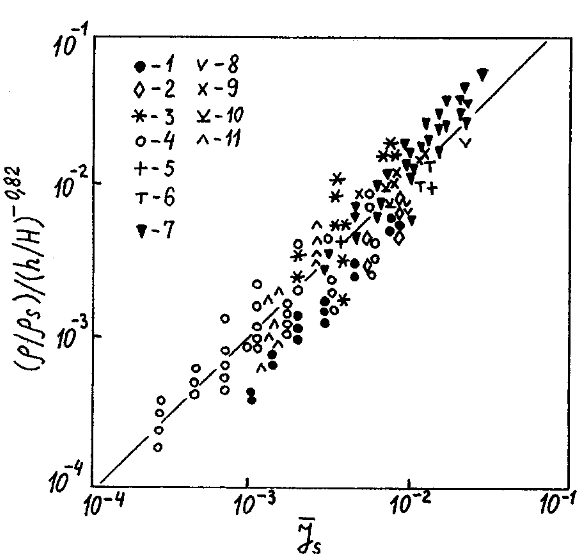

Fig. 8. Generalization of experimental data on particle concentration in the Catalytic Cracking CFB riser: 1, [9] (d = 0.287 mm); 2, [10] (0.049); 3, [11] (0.046); 4, [12] (0.078); 5, 6, [13] (0.087, 0.227); 7, [14] (0.061); 8, [15] (0.076); 9, [16] (0.085); 10, [17] (0.064); 11, [18] (0.240).

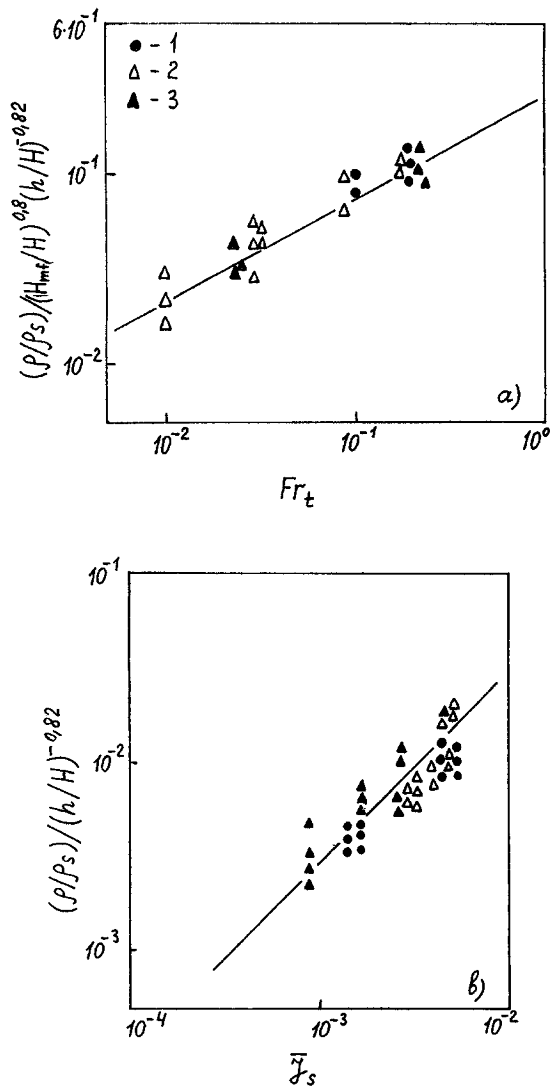

Fig. 9. Generalization of experimental data on particle concentration in the CFB Combustor riser: 1, sand d = 0.250 mm; 2, anionite d = 0.550 mm; 3, anionite d = 0.650 mm.

on the basis of the functional relation (21) by the least-square method. The obtained correlation has the form:

$$Nu_{c\text{-}c} = 55.0\left(\frac{h}{H}\right)^{-0.42} Ar^{0.12}\bar{J}_s^{0.50}\left(\frac{d}{L}\right)^{0.43}. \tag{27}$$

The experimental points and relation (27) are shown in Fig. 10. A root-mean-square deviation of calculated values of $\alpha_{c\text{-}c}$ from experimental values is 17%. The range of verification of (27) is $5.91 \leq J_s \leq 68.7$ kg/m^2

s; $6.65 < H \le 10$ m; $0.046 \le d \le 0.287$ mm; $0.07 \le L \le 0.4$ m; $50 \le Ar \le 2200$.

The obtained Eqs. (23) and (27) make it possible to find a simple functional relation between $\alpha_{c\text{-}c}$ and ρ. A combination of (23) and (27) leads to the relation

$$Nu_{c\text{-}c} = 55.0\left(\frac{\rho}{\rho_s}\right)^{0.50} Ar^{0.12}\left(\frac{d}{L}\right)^{0.43}, \tag{28}$$

which already has a universal character with respect to the scheme of CFB operation. The numbers $\bar{J}_s$, Fr_t, and H_{mf}/H are not present in this relation explicitly. By virtue of this fact, the data obtained in CFB operating according to different schemes are processed by (28). Fig. 11 shows the results of comparison of the data of [13,20–28] with those calculated by (28).

A fairly good qualitative agreement of calculated and experimental values of $Nu_{c\text{-}c}$ is observed when $d/L < 10^{-3}$. At larger values of d/L experimental data turn to be lower than those calculated and the dependence on this simplex degenerates. This case corresponds to the conditions of local heat transfer (small lengths of a heat transfer surface) when a heat transfer rate is maximum due to small times of contact between a disperse medium and the surface, and is limited only by thermal resistance of a gas film separating particles of the first row from the heat transfer surface.

As is seen from Fig. 11, under the conditions of local heat transfer a rather substantial scatter of experimental data obtained by different researchers is observed. This can be explained by the dependence of ρ_w/ρ (or the profile of particle concentration in a horizontal section of the riser) on the riser diameter [29]. The local coefficient $\alpha_{c\text{-}c} = \alpha_f$ (a limiting value of $\alpha_{c\text{-}c}$) is determined by a gas film thickness. Since the thickness of this film depends on ρ_w, then the coefficient α_f will be determined by a value of ρ at the heat transfer surface, i.e. by a value of ρ_w. A comparison of different experimental data was made at equal values of mean concentration ρ and this, probably, led to their scattering because the data are obtained in different installations in which the ratio ρ_w/ρ is not constant, but depends on the riser diameter. By virtue of this, experimental data on local heat transfer were generalized independently by the dependence

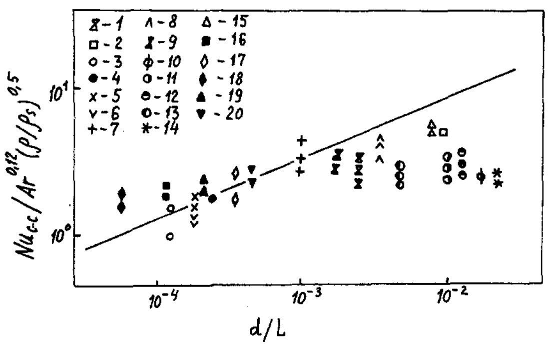

Fig. 11. Dependence of the conductive–convective heat transfer coefficient on dimensionless concentration of particles (generalization of experimental data): 1, [21] (d = 0.250 mm); 2, [13] (0.227); 3, 4, [24] (0.188, 0.356); 5, 6, [23] (0.241); 7, [20] (0.309); 8, [13] (0.087); 9, [21] (0.170); 10, [22] (0.125); 11–13, [present study] (0.550, 0.650, 0.250); 14, [25] (0.240); 15, [26] (0.171); 16, 17, [27] (0.058, 0.165, P = 1 bar); 18–20, [28] (0.04, 0.13, 0.30).

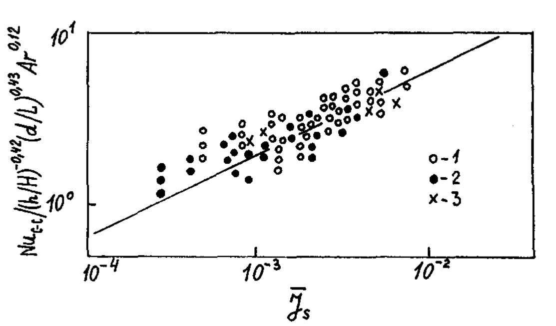

Fig. 10. Generalization of experimental data on conductive–convective heat transfer in Catalytic Cracking CFB: 1, [9] (d = 0.287 mm); 2, [12] (0.078); 3, [11] (0.046).

$$Nu_f = 240\left(\frac{\rho}{\rho_s}\right)^{0.50}\left(\frac{d}{D}\right)^{0.50}, \tag{29}$$

which explicitly allows for the effect of the riser diameter (and so a real profile of particle concentration in the horizontal section of the riser) (Fig. 12). The range of verification of (29) is: $0.171 \le d \le 0.65$ mm; $0.152 \le D \le 0.7$ m; $10 \le \rho \le 140$ kg/m^3.

To determine the region of the applicability of Eq. (29) more accurately an estimate of limiting values of L should be obtained, when it is already admissible to consider heat transfer to be local. In [30], on the basis of the analysis of nonstationary heat exchange of disperse medium with the surface, the author established that at small contact times ($Fo = a_f t/d^2 \le 5$) heat transfer virtually stops to depend on time, thus reaching limiting values α_f. Using this result it is easy to obtain an order of a minimum value of L:

$$L_{min} \approx 10^5 d^2, \tag{30}$$

where L_{min} and d are expressed in meters. Here the relations $t = L/v$, v = 1 m/s [11]; $a_f = 2.25 \times 10^{-5}$ m^2/s were used.

4.2.2. CFB Combustor

Using the earlier obtained correlation (24) on the basis of (29) the expression for calculating local heat transfer in the form of (22) is

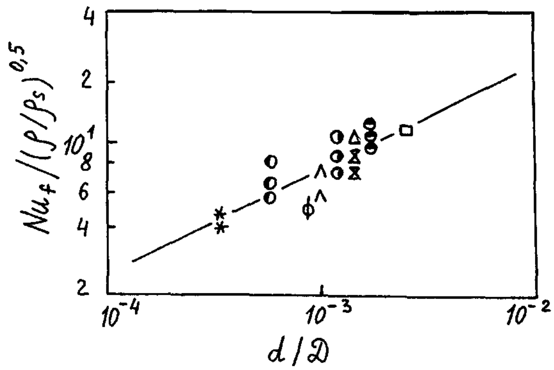

Fig. 12. Local heat transfer in a CFB (limiting value of α_{c-c}). For notation see Fig. 11.

$$Nu_f = 122.4\left(\frac{h}{H}\right)^{-0.41} Fr_t^{0.27}\left(\frac{d}{D}\right)^{0.50}\left(\frac{H_{mf}}{H}\right)^{0.40}. \quad (31)$$

With account for very limited range of checking, (31) requires further refinement.

5. Conclusion

Within the framework of the similarity theory, the rules of scaling transfer processes in CFB operating were formulated according to the schemes Catalytic Cracking CFB and CFB Combustor.

Complete systems of dimensionless groups determining similarity of circulating fluidized beds, operating by the schemes Catalytic Cracking CFB (8) and CFB Combustor (13), are found.

Particle concentration and the local heat transfer coefficient on a cold model of a boiler with a CFB are studied experimentally. The dependence of these values on the determining factors is revealed.

On the basis of the developed method of scaling the obtained, and literature data on distribution of particle concentration (23)–(25) and on the conductive–convective heat transfer coefficient (27) and (28) over the height of the transport zone of the riser, was generalized.

The expression for local (limiting) values of α_{c-c} (29) is obtained.

A fruitful character of use of numbers $\bar{J}_s$ and Fr_t that are important generalized characteristics of Catalytic Cracking CFB and CFB Combustor, respectively, is shown.

Systems of dimensionless groups determining similarity of transfer processes with CFB of different types make it possible to model industrial installations with a CFB, give methodology for analysing the results obtained and for obtaining wide generalizations of experimental data in a convenient dimensionless form.

Acknowledgement

One of the authors (Yu.S.T.) would like to express his gratitude to the INTAS Foundation for financial support of investigations on similarity of circulating fluidized beds.

References

[1] M. Horio, H. Ishii, Y. Kobukai, N. Yamanishi, A scaling law for circulating fluidized beds, J. Chem. Engng Japan 22 (1989) 587–592.

[2] E.H. Van der Meer. The effect of exit geometry on flow in a circulating fluidized bed. CPGS dissertation, University of Cambridge, 1994.

[3] L.R. Glicksman, M. Hyre, K. Woloshun, Simplified scaling relationships for fluidized beds, Powder Tech. 77 (1993) 177–199.

[4] L.R. Glicksman, D. Westphalen, C. Brereton, J. Grace, Verification of scaling laws for circulating fluidized beds, in: Proceedings of the Third International Conference on CFB Nagoya, Japan, 1998, pp. 119–124.

[5] H. Chang, M. Louge, Fluid dynamic similarity of circulating fluidized beds, Powder Tech. 70 (1992) 259–270.

[6] B. Leckner, F. Johnsson, The aspect ratio—an important parameter for comparison of the fluid mechanics of CFB risers. Workshop 'Modelling and Control of Fluidized Bed Systems'. Technische Universität Hamburg–Harburg, May 22–23, 1995.

[7] H.E. Huntley, Dimensional Analysis, New York, 1967.

[8] V.A. Borodulya, Yu.S. Teplitskiy, Scaling in circulating fluidized beds, Proc. Minsk Int. Forum on Heat and Mass Transfer Minsk, Belarus 5 (1996) 69–74.

[9] A.P. Baskakov, Flow through fluidized system, in: V.G. Anstein, A.P. Baskakov (Eds.), Fluidization, Khimiya, Moscow, 1991, pp. 333–375.

[10] V.K. Maskaev, A.P. Baskakov, A.G. Usoltsev, Aerodynamics of circulating fluidized bed, Inzh-Fiz. Zh. 67 (1989) 762–767.

[11] W. Nowak, A comprehensive study of the circulating fluidized bed. A thesis presented to qualify as Assistant Professor Institute of Heat Machinery, Czestohowa, 1989.

[12] A.P. Baskakov, V.K. Maskaev, I.V. Ivanov, A.G. Usoltsev, Influence of hydrodynamics of circulating fluidized bed on the intensity of external heat transfer, Inzh-Fiz. Zh. 61 (1991) 778–781.

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

[14] K. Kato, H. Shibasaki, K. Tamura, S. Arita, C. Wang, T. Takarada, Particle holdup in a fast fluidized bed, J. Chem. Engng Japan 22 (1989) 130–136.

[15] R. Bader, J. Findlay, T.M. Knowlton, Gas/solids flow patterns in a 30.5-cm diameter circulating fluidized bed, in: Proceedings of the Second International Conference on CFBs, Compiegne, 1988, pp. 123–138.

[16] E.-U. Hartge, D. Rensner, J. Werther, Solid concentration and velocity patterns in circulating fluidized

beds, in: Proceedings of the Second International Conference on CFBs, Compiegne, 1988, pp. 165–180.

[17] M.J. Rhodes, D. Geldart, The upward flow of gas/solid suspensions, Part 1, Chem. Engng Des. 67 (1989) 20–29.

[18] V.A. Borodulya, Yu.S. Teplitskiy, Yu.G. Epanov, Regular patterns of local heat transfer in circulating fluidized bed, Proc. 1st Russian Nat. Heat Transfer Conf. 7 (1994) 34–38.

[19] F. Johnsson, B. Leckner, Vertical distribution of solids in a CFB-furnace, in: Proceedings of the Thirteenth International Conference on Fluidized Bed Combustion, Orlando, 1995.

[20] P.K. Nag, M.S. Ali, Effect of operating parameters on bed-to-wall heat transfer in a high-temperature circulating fluidized bed, Int. J. Energy Res. 16 (1992) 61–73.

[21] H. Kobro, C. Brereton, Control and fuel flexibility of circulating fluidized beds, in: P. Basu (Ed.), Circulating Fluidized Bed Technology, Pergamon Press, Toronto, 1986, pp. 263–272.

[22] S. Dou, B. Herb, K. Tuzla, J.C. Chen, Dynamic variation of solid concentration and heat transfer coefficient at wall of circulating fluidized bed, in: Proceedings of the International Conference 'Fluidization-VII', 1992, pp. 793–801.

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

[24] R.L. Wu, C.J. Lim, J. Chaouki, J.R. Grace, Heat transfer from a circulating fluidized bed to membrane waterwall surfaces, AIChE J. 33 (1987) 1888–1893.

[25] F. Johnsson, B.-A. Andersson, B. Leckner, Heat transfer in circulating fluidized boilers, Proc. 1st Minsk Int. Forum on Heat and Mass Transfer 4 (5) (1989) 185–195.

[26] R.L. Wu, C.J. Lim, J.R. Grace, The measurement of instantaneous local heat transfer coefficients in a circulating fluidized bed, Can. J. Chem. Engng 67 (1989) 301–307.

[27] K.-E. Wirth, Prediction of heat transfer in circulating fluidized beds, in: Proceedings of the Fourth International Conference on Circulating Fluidized Beds, Pittsburg, 1993, pp. 344–349.

[28] D. Shi, L. Reh, Wall-to-bed heat transfer in circulating fluidized beds, Proc. 5th Int. Conf. on Circulating Fluidized Beds, Beijing HM6 (1966) 1–6.

[29] L.R. Glicksman, Circulating fluidized bed heat transfer, in: Proceedings of the Second International Conference on CFBs, Compiegne 1988, 1988, pp. 13–29.

[30] Yu.S. Teplitskiy, Hydrodynamics and heat transfer in free and packed fluidized beds. Doctor thesis. Institute of Thermo-physics, Novosibirsk, 1990.