A SEMI-EMPIRICAL MODEL TO CALCULATE THE INERT SOLIDS INVENTORY AND THE MAIN DIMENSIONS OF A CFB REACTOR AIMING BIOMASS THERMOCHEMICAL CONVERSION

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Abstract. *The inert solids inventory is a recognized key parameter in designing circulating fluidized bed (CFB) reactors. In such systems, the total mass (solids holdup) of particles in the loop has a special role in guaranteeing the operational stability of thermochemical processes for heat and electricity generation involving solid fuels with low ash fraction, like biomass and some wastes. In this work, a practical step-by-step procedure is proposed to determine both the inert solids inventory and the main loop dimensions of a bench-scale circulating fluidized bed reactor. The model is based on semi-empirical hydrodynamic correlations, the pressure balance principle, and useful data obtained from previous works of the literature. In simulations, quartz sand with Sauter diameter of 200 µm and air at atmospheric pressure are used as inert material of the bed and fluidizing agent, respectively. The mean temperature of the CFB system loop is assumed to be 800°C in order to simulate either combustion or gasification conditions. Results are successfully compared with experimental data reported by other authors.*

Keywords: *circulating fluidized bed, biomass thermochemical conversion, solids holdup.*

1. INTRODUCTION

Circulating fluidized beds (CFBs) are gas-solid contactor systems where fine solids particles are transported in a long vertical duct (riser) by a high velocity gas stream. After leaving the top of the column, the solids are usually separated from the gas and circulated again to the base of the riser by using an injection valve. Normally, risers work in the fast fluidization regime. In this regime, the gas superficial velocity into the column overcomes the solids transport velocity (Bai *et al.*, 1993). Chemical reacting systems requiring high specific transfer rates, high solids throughput and thermal uniformity within the reactor are excellent candidates for the use of CFB technology (Hartge *et al.*, 1986). Particularly, the combustion of low grade fossil fuels, biomass and waste for electricity production under strict environment control represents one of the most successful applications of CFB systems (Berruti and Kalogerakis, 1989). In last years, industrial processes such as low temperature absorption (dry scrubbing), biomass pyrolysis and gasification, Fischer-Tropsch synthesis and other catalityc and non-catalytic reactions have also been studied in CFB reactors (Xianwen, *et al.*, 2000; Barišić, *et al.*, 2005; Suraniti, *et al*., 2009).

The bed solids inventory is an important parameter for the successful design and operation of a CFB reactor. Together with the superficial gas velocity and the solids circulation flux it drives the bed particles distribution in the CFB loop system, which is composed by the riser, cyclone, standpipe and solids circulation valve. The solids distribution governs the pressure drop along the system and the particle residence time within the fast bed zone, where the main chemical reactions take place. Moreover, the solids distribution shows strong effects on the mass-transfer rate and heat transfer coefficients in the CFB loop (Adánez, *et al.*, 1994). For combustion and gasification applications, the inert solids used in the system represent more than 97% in mass of the total solids inventory (Basu, 2006). In most of cases, the inert solid material is quartz sand because of its relative low cost and excellent performance at high temperatures.

Many authors have studied the hydrodynamics of fast bed systems, and nowadays, the gas-solid behavior of a CFB loop in terms of its geometrical configuration, particles properties in the bed, solids inventory, superficial gas velocity and solids circulation flux is relatively well known (Rhodes and Laussmann, 1992; Mastellone and Arena, 1999; Kim and Kim, 2002; Qi *et al.*, 2008). Several models based on both theoretical analysis and experimental tests have been proposed in the past for explaining the operation of CFB systems or some of its components at high or environmental temperature (Knowlton, 1988; Wang *et al.*, 1996; Davidson, 2000; Tong, *et al.*, 2003; Gungor and Eskin, 2008). However, for design proposal a few works describing straight procedures to define the operational characteristics and geometry of the components of a CFB loop are reported in the literature (Chong, *et al.*, 1988; Yin, *et al.*, 2002, Dewil, *et al.*, 2008, Ramírez *et al.*, 2009).

This work presents a practical hydrodynamic model to determine the bed solids inventory and other basic operational-dimensional parameters of a bench-scale CFB system for future application in biomass thermochemical conversion. The procedure is based on semi-empirical correlations and data obtained from the previous works. Only some physical properties of both the inert material in the bed and the fluidization gas, as well as, the inner diameters of the fast bed column of the loop are needed as input data. Details about the procedure are shown at follows.

2. THE SEMI-EMPIRICAL MODEL

For convenience, the procedure of the semi-empirical model is described in three stages, one for each main component of the CFB system (riser, cyclone, and standpipe-solid recycle valve). In it, the main loop dimensions and operational conditions are initially defined. After that, the solids inventory required for stable operation of the loop is calculated.

2.1 The Riser

The riser height H can be known from the empirical entrainment model developed by Kunii and Levenspiel (1990). According this model, H is related to the respective local axial voidage fraction in the bed ε_H by the exponential expression:

$$\frac{\varepsilon_d - \varepsilon_H}{\varepsilon_d - \varepsilon_a} = exp\left[-a\left(H - h_i\right)\right] \tag{1}$$

where h_i corresponds to the inflexion point of the characteristic "S" voidage profile found in fast bed regimes. For combustion applications, h_i is normally the level of the secondary air-injection, which is measured vertically from the riser base (Basu, 2006). This author also indicates that the asymptotic voidage value in the denser section (bottom of the riser) ε_a is between 0,78 and 0,88 for fast beds using Group A particles of the Geldart classification, being it a little higher when particles of Group B are employed.

For use in Eq. (1), Kunni and Levenspiel (1991) determined the behavior of the decay constant a in terms of the superficial gas velocity U and the sieve diameter of the bed particles d_s. From the experimental data, a can be approximated by the Eq. (2):

$$a = \frac{4.1011 ln\left(d_s.10^6\right) - 15.181}{U} \tag{2}$$

According to Yang (1983), no noticeable deviation is obtained when the asymptotic voidage in the dilute phase (upper section of the riser) ε_d is adopted as the choking voidage ε_{ch}, which is defined implicitly by the Eq. (3). Although this equation is validated for internal riser diameters D minor than 0.3 m, environmental temperature and pressure conditions, and Group A particles, it can be used as a first approximation for others cases and operation conditions.

$$\frac{U_{ch}}{\varepsilon_{ch}} = U_{t,ns} + \sqrt{\frac{2gD\left(\varepsilon_{ch}^{-4.7} - 1\right)\rho_s^{2.2}}{6.81x10^5 \rho_g^{2.2}}} \tag{3}$$

In Eq. (3), g is the acceleration due to gravity, U_{ch}, the choking velocity, ρ_s and ρ_g, the bed particle and the fluidizing gas density, respectively. Additionally, the terminal velocity for single particles U_t and the choking velocity of the CFB system can be associated to the solids circulation flux at the riser exit G_s by:

$$G_s = \left(U_{ch} - U_t\right)\left(1 - \varepsilon_{ch}\right)\rho_s \tag{4}$$

where U_t is determined through the expressions:

$$\begin{aligned}
&\frac{d_s U_t \rho_g}{\mu} = \frac{Ar}{18} && \text{for} \quad Re < 0.4 \\
&\frac{d_s U_t \rho_g}{\mu} = \left(\frac{Ar}{7.5}\right)^{0.666} && \text{for} \quad 0.4 < Re < 500 \\
&\frac{d_s U_t \rho_g}{\mu} = \left(\frac{Ar}{0.33}\right)^{0.5} && \text{for} \quad 500 < Re
\end{aligned} \tag{5}$$

On the other hand, the solids terminal velocity for non-spherical particles $U_{t,ns}$ can be calculated from the Eq. (6), where U_t is multiplied by the factor $K_{t,ns}$ defined in Eq. (7) (Basu, 2006).

$$U_{t,ns} = K_{t,ns} U_t \tag{6}$$

with,

$$K_{t,ns} = 0.843 log_{10}\left[\frac{\phi_s}{0.065}\right] \quad \text{for} \quad \left\{Re = \frac{d_s U_t \rho_g}{\mu}\right\} < 0.2$$

$$K_{t,ns} = \left[\frac{4.89(\rho_s - \rho_g) g d_s}{3\rho_g (5.31 - 4.88\phi_s)}\right]^{0.5} \quad \text{for} \quad \left\{Re = \frac{d_s U_t \rho_g}{\mu}\right\} > 1{,}000 \tag{7}$$

where ϕ_s is the particle sphericity, and μ, the gas viscosity. For $0.2<Re<1{,}000$, $K_{t,ns}$ is calculated by interpolation.

In the solution procedure, the superficial gas velocity U is settled equal to both the transport velocity of the particles U_{tr} and the choking velocity U_{ch} in order to guarantee the minimum condition for operation stability of a fast fluidized bed (Adánez *et al.,* 1993). Several relations for U_{tr} are known in the literature. In this work, the expression proposed by Perales *et al.* (1991) is used:

$$U_{tr} = U_{ch} = U = 1.45\left(\frac{\mu}{\rho_g d_s}\right) Ar^{0.484} \quad \text{for } 20 < Ar < 50{,}000 \tag{8}$$

Alternatively, the axial voidage fraction ε_H can be known through the expression suggested by Davidson (2000):

$$(1-\varepsilon_H)\rho_s = \left(\frac{E}{U} + \frac{w}{U_f}\right) \tag{9}$$

with,

$$G_s = E - w \tag{10}$$

In Eq. (9), U_f is the falling velocity of particles at the riser wall. The solid flux moving downward w and the solid flux moving upward E in Eq. (10) are related by the internal solids refluxing ratio $R_s = w/E$. Therefore, if D, h_i, R_s, ε_a and the properties of the solids and fluidizing gas are given, the Eqs. (1) to (10) are used to find the values of H, G_s, U and ε_H by iteration technique.

Finally, according to Adánez *et al.* (1994), the solids inventory in the fast bed column $I_{s,r}$ and the respective pressure drop ΔP_r are calculate as:

$$I_{s,r} = \left(\frac{\pi D^2 \rho_s}{4}\right)\left[\left(\frac{\varepsilon_H - \varepsilon_a}{a}\right) + H(1-\varepsilon_a) - (H - h_i)(\varepsilon_{ch} - \varepsilon_a)\right] \tag{11}$$

$$\Delta P_r = \frac{4 g I_{s,r}}{\pi D^2} + \frac{G_s^2}{\rho_s (1-\varepsilon_H)} \tag{12}$$

2.2 The Cyclone

For this work, a high efficiency tangential cyclone with the Swift geometrical configuration was considered as the particle control device included in the CFB loop (Figure 1). According to Basu (2006), the geometrical relationships are:

$$A = 0.44D_c; \quad C = 0.21D_c; \quad M = 0.4D_c; \quad F = 0.5D_c; \quad S = 1.4D_c; \quad B = 3.9D_c; \quad N = 0.4D_c \tag{13}$$

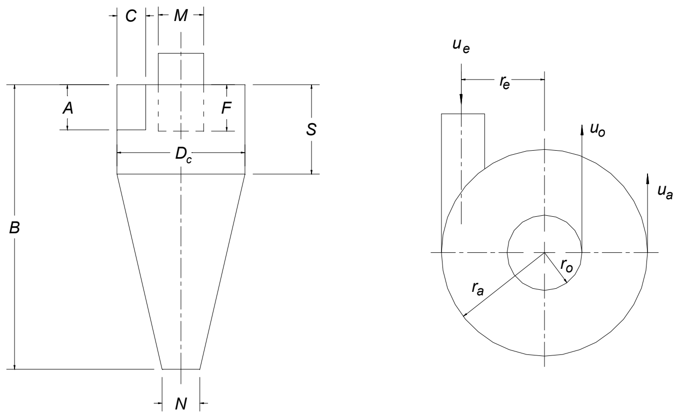

Figure 1. Geometrical configuration of the cyclone.

The internal diameter of the cylindrical section D_c is expressed in function of the gas flow rate entering into the cyclone Q by Eq. (14),

$$D_c = \sqrt{\frac{Q}{1.37}} \tag{14}$$

On the other hand, the total pressure drop in the cyclone ΔP_c is calculated through the model suggested by Muschelknautz and Greif (1997):

$$\Delta P_c = f_w \frac{A_R}{V_b} \frac{\rho_g}{2} \left(u_a u_o\right)^{1.5} + \left[2 + 3\left(\frac{u_o}{v_o}\right)^{4/3} + \left(\frac{u_o}{v_o}\right)^2\right] \frac{\rho_g}{2} v_o^2 \tag{15}$$

where the gas flowing through cyclone barrel V_b is $0.9(\pi D^2/4)U$. The total wall friction coefficient due to the solid-gas suspension f_w is approximated by the Eq. (16), which is function of both the solid to gas mass ratio entering to the cyclone C_e and the wall friction coefficient for clean gas flow f_0 (Basu, 2006):

$$f_w = f_0\left(1 + 2\sqrt{C_e}\right) \tag{16}$$

In this work, the C_e value was obtained from the term G_s/U. The value of f_0 can be adopted as 0.005 for Reynolds numbers higher than 1,000 (expected for CFB systems) and pipe wall relative roughness in the range of 2.6×10^{-5} to 6×10^{-4} (Basu, 2006). Additionally, according to the model of Muschelknautz and Greif (1997):

$$u_o = \frac{u_a r_o}{r_o + \frac{f_w}{2} \frac{4A_R}{\left(\pi D^2 U\right)} u_a \sqrt{r_a r_o}} \tag{17}$$

$$u_a = \frac{u_e r_e}{r_a \alpha} \tag{18}$$

$$\alpha = \frac{1}{\beta}\left[1 - \sqrt{1 - 4\left[\frac{\beta}{2} - \left(\frac{\beta}{2}\right)^2\right]} \sqrt{1 - \left(\frac{1-\beta^2}{1+C_e}\right)\left(2\beta - \beta^2\right)}\right] \tag{19}$$

$$\beta = \frac{C}{r_a} \tag{20}$$

where u_e is the gas velocity in the inlet section of the cyclone (Figure 1). Additionally, u_o and u_a are the tangential gas velocity at cyclone barrel and the gas exit tube radius, respectively. The term A_R is the wall area of the cyclone, including roof and outer surface of the outlet tube. Finally, the mean gas velocity within the exit duct of the cyclone v_o is calculated as:

$$v_o = \frac{U}{2\pi r_o} \tag{21}$$

2.3 The Standpipe and the L-Valve

The standpipe is a vertical tube with diameter normally constant that receives the solids discharged by the cyclone. In CFB systems operating with L-valve, the length of the standpipe is measured from the cyclone bottom to the gas injection level of the L-valve vertical leg. Initially, the selected internal diameter for both the standpipe and the L-valve is that obtained for the solids exit tube of the cyclone. In the proposed model, the standpipe internal diameter is only adopted after verifying that the solids velocity value in the moving bed at the bottom of the standpipe $U_{s,lv}$ is minor than 0.15 m/s (Knowlton, 1988). This velocity is calculated dividing the standpipe solids circulating flow by the packed bed solids density.

On the other hand, if the L-valve is operated at minimum fluidization conditions, the height of the solids above of the aeration point $L_{v,lv}$ can be determined through the expression suggested by Knowlton (1997):

$$L_{v,lv} = \frac{\Delta P_{lv} + \Delta P_{r,sr} + \Delta P_c}{\left(\frac{\Delta P}{L}\right)_{mf,sp}} \tag{22}$$

According to the pressure balance around the CFB loop, the numerator of the Eq. (22) represents the pressure drop in the standpipe ΔP_{sp}. On the other hand, the denominator characterizes the pressure drop due to the height of the solid bed at minimum fluidization condition in the vertical leg of the L-valve, which is defined as:

$$\left(\frac{\Delta P}{L}\right)_{mf,sp} = \rho_s g\left(1-\varepsilon_{mf}\right) \tag{23}$$

with ε_{mf} being the voidage fraction at minimum fluidization condition for the bed particles. For the proposed model, the same voidage fraction is assumed in the horizontal section of the L-valve.

In Eq.(22), ΔP_{lv} is the pressure drop in the L-valve calculated from the aeration point to the solids discharge position into the riser. Geldart and Jones (1991) obtained the Eq. (24) from experimental data to calculate ΔP_{lv} in function of the L-valve internal diameter D_{lv}, the sieve diameter of the bed particles d_s, and the solids circulation flux crossing the injection device $G_{s,lv}$.

$$\Delta P_{lv} = 216\left(\frac{G_{s,lv}^{0.17}}{D_{lv}^{0.63} d_s^{0.15}}\right) \tag{24}$$

Because of the standpipe and the L-valve have the same internal diameter, the solids circulation flux in the return leg is a constant value that can be determined by mass balance from the G_s and the riser to standpipe area ratio.

Also in Eq. (22), $\Delta P_{r,sr}$ is the pressure drop in the vertical section of the riser measured above the solids return level. It can be found by difference from the pressure drop in the total riser height:

$$\Delta P_{r,sr} = \Delta P_r - h_{sr}\left(1-\varepsilon_a\right) g \rho_s \tag{25}$$

where h_{sr} is the solids return level measured above of the riser base. In the proposed model, the value of the h_{sr} was constrained to $h_{sr} \leq 0.5 h_i$.

In order to find the standpipe and L-valve main dimensions, the model considers the recommendations given by Knowlton (1988):

- Standpipes are typically designed to be 1.5 to 2 times $L_{v,lv}$. In this work, the actual length of the standpipe was chosen to guarantee at least $2L_{v,lv}$ in order to maximize the residence time of the recycled solids, which is always important in combustion and gasification processes;

- The length of the horizontal section of the L-valve $L_{h,lv}$ should be in the range of 1.5 to 10 times its internal diameter. In the model, the mean value of the range above was adopted;

- The aeration point level L_a should be localized above the center line of the horizontal tube at approximately two times the internal diameter of the L-valve.

On the other hand, the external aeration mass flow rate required by the L-valve operation, m_{lv} is calculated according to the experimental equation presented by Geldart and Jones (1991):

$$m_{lv} = m_{mf}\left[\frac{\left(\frac{G_{s,lv}}{D_{lv}}\right) + 2{,}965}{3{,}354}\right] \tag{26}$$

where, m_{mf} is the mass flow rate of the aeration gas needed for minimum fluidization condition of the bed particles within the standpipe.

Finally, the solids inventory in the standpipe and the L-valve section is calculated as:

$$I_{s,sp-lv} = \frac{\left(L_{v,lv} + L_{h,lv}\right)\left(1-\varepsilon_{mf}\right)\left(\pi D_{lv}^2\right)\rho_s}{4} \tag{27}$$

Therefore, considering that the particles of the CFB system are mainly concentrated in the riser, the standpipe and the L-valve, the total solids inventory is:

$$I_{s,T} = I_{s,r} + I_{s,sp-lv} \tag{28}$$

The Eqs. (22) to (28) are solved by using an iterative procedure, considering the geometrical restrictions previously specified for the cyclone, the standpipe and the L-valve. Figure 2 schematizes the structure of the semi-empirical model described above.

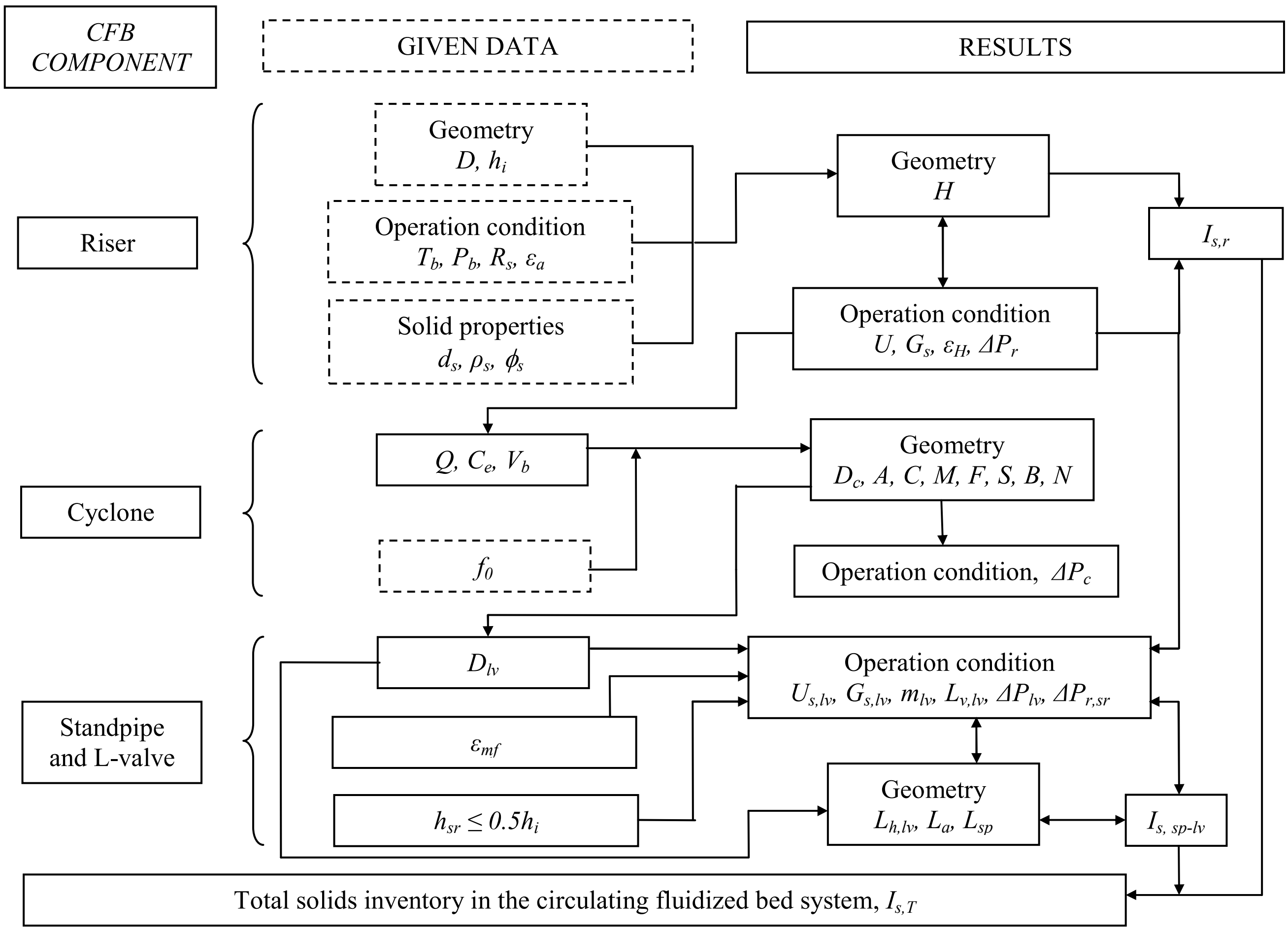

Figure 2. Structure of the semi-empirical model (T_b and P_b are the temperature and pressure of the system, respectively).

3. RESULTS AND DISCUSSION

The semi-empirical model was used to determine the main dimensions and operational conditions of a bench-scale CFB loop, looking forward to ease the preliminary system setup for biomass conversion via thermochemical processes, such as combustion or gasification. The input data used for the simulation for this design case and its respective results are presented in Tab. 1 and Tab. 2, respectively. Quartz sand is used as the bed material within the CFB system.

Table 1. Input data used in the semi-empirical model (design case)

Parameter	Value	Reference
Internal diameter of the riser column, D [m]	0.074	Assumed.
Height of the secondary air injection, h_i [m]	0.8	Assumed.
Solids refluxing ratio, R_s [-]	0.3	Smolders and Baeyens, (2000); Davidson (2000).
Cluster velocity, U_f [m/s]	1.2	Grace (1990); Davidson (2000).
Voidage fraction in the bottom section of the riser, ε_a [-]	0.9	Basu (2006).
Operation temperature, T_b [°C]	800	Assumed.
Operation pressure, P_b [kPa]	101.3	Assumed.
Sieve diameter of the bed particles, d_s [m]	200×10^{-3}	Assumed.
Solids density, ρ_s [kg/m^3]	2,650	Perales, *et al.* (1991).
Solids sphericity, ϕ_s [-]	0.75	Basu (2006).
Voidage fraction at minimum fluidization, ε_{mf} [-]	0.45	Arena *et al.* (1998).

Table 2. Results obtained from the semi-empirical model (design case)

Component of the CFB system	Operational/Dimensional Parameter	Value
Riser	Particle terminal velocity corrected by sphericity, $U_{t,ns}$ [m/s]	1.70
	Superficial gas velocity, $U = U_{tr} = U_{ch}$ [m/s]	5.45
	Solids circulation flux in the riser, G_s [kg/m^2.s]	35.71
	Choking voidage, ε_{ch} [-]	0.9964
	Voidage fraction at the riser exit, ε_H [-]	0.9901
	Riser height, H [m]	3.07
	Riser height to internal diameter ratio, R_{H-D} [-]	41.46
	Solids inventory, $I_{s,r}$ [kg]	1.86
	Pressure drop in the total riser height, ΔP_r [Pa]	4,287
	Solids return level above the riser base [m]	0.4
	Pressure drop in the vertical section of the riser measured above the solids return level, $\Delta P_{r,sr}$ [Pa]	3,248
Cyclone	Volumetric gas flow entering to cyclone, Q [m/s]	0.023
	Solid to gas mass ratio entering to cyclone, C_e [-]	19.9
	Gas velocity at the inlet section of the cyclone, u_e [m/s]	14.83
	Cyclone dimensions [m]: $Dc = 0.131$; $A = 0.058$; $C = 0.027$; $M = 0.052$; $F = 0.065$; $S = 0.183$; $B = 0.511$; $N = 0.052$.	-
	Gas velocity at cyclone exit, v_o [m/s]	10.90
	Pressure drop in the cyclone, ΔP_c [Pa]	150
Standpipe and L-valve	Internal diameter of the standpipe and the L-valve, D_{lv} [m]	0.052
	Solids circulation flux in the standpipe and the L-valve, $G_{s,lv}$ [kg/m^2.s]	72.3
	Solids velocity in the moving bed, $U_{s,lv}$ [m/s]	0.05
	Horizontal section of the L-valve $L_{h,lv}$ [m]	0.30
	Aeration level above of the horizontal section center line, L_a [m]	0.104
	Aeration mass flow rate in the L-valve, m_{lv} [kg/h]	0.50
	Height of the solids above of the aeration point, $L_{v,lv}$ [m]	0.53
	Height of the standpipe, L_{sp} [m]	2.05
	Pressure drop in the L-valve, ΔP_{lv} [Pa]	4,165
	Pressure drop in the standpipe, ΔP_{sp} [Pa]	7,563
Total solids inventory in the CFB system [kg]		4.83

Experimental data for straight comparison of the results obtained above are unavailable. However, a rough analysis from previous works found in literature (Patience and Chaouki, 1993; Arena *et al.*, 1998; Mastellone and Arena, 1999) shows that the scaling parameter $[U/(gD)^{0.5}]^{-0.3}Gs/\rho_sU$, suggested by Qi *et al.* (2008), has a mean deviation of around 11%. Additionally, the riser height to internal diameter ratio found in the simulation was about 36% lower. When the total solids inventory to total volume of the CFB loop ratio was considered, a mean deviation of approximately 6% was obtained (Arena *et al.*, 1998; García-Ibañez *et al.*, 2004).

In order to verify more rigorously the validity of the model, results of simulation were compared with experimental data obtained by Hory *et al.* (2006) during the operation of a bench-scale CFB coal combustor tested under steady state pre-heating condition. In such work, a fast fluidized bed of quartz sand was maintained constant at approximately 400ºC by previously heated fluidizing air fed at the riser bottom. In the simulation, the internal solids refluxing ratio R_s was settled at the appropriated value to guarantee the same riser height of the experimental CFB combustor. Table 3 shows the input data used by the model and the experimental tests for comparison.

Table 3. Input experimental data used in the model (Hory *et al.*, 2006)[1].

Parameter	Value
Internal diameter of the riser column, D [m]	0.102
Height of the secondary air injection, h_i [m]	0.9
Solids refluxing ratio, R_s [-]	0.066
Operation temperature, T_b [°C]	400
Sieve diameter of the bed particles, d_s [m]	353x10^{-3}
Solids density, ρ_s [kg/m^3]	2,700

[1] Input data values no specified were assumed equal to those presented in Tab. 1.

The main results obtained from the simulation are presented in Tab. 4. Additionally, the deviations from experimental tests are illustrated in Fig.3.

Table 4. Results obtained from the semi-empirical model (Hory *et al*, 2006 - Comparison case)

Component of the CFB system	Operational/Dimensional Parameter	Value
Riser	Particle terminal velocity corrected by sphericity, $U_{t,ns}$ [m/s]	3.04
	Superficial gas velocity, $U = U_{tr} = U_{ch}$ [m/s]	5.54
	Solids circulation flux in the riser, G_s [kg/m^2.s]	20.93
	Choking voidage, ε_{ch} [-]	0.9969
	Voidage fraction at the riser exit, ε_H [-]	0.9962
	Riser height, H [m]	4.00
	Riser height to internal diameter ratio, $R_{H\text{-}D}$ [-]	39.2
	Solids inventory in the total riser height, $I_{s,r}$ [kg]	3.52
	Pressure drop in the total riser height, ΔP_r [Pa]	4,268
	Solids return level above the riser base [m]	0.45
	Pressure drop in the vertical section of the riser measured above the solids return level, $\Delta P_{r,sr}$ [Pa]	3,077
Cyclone	Volumetric gas flow entering to cyclone, Q [m/s]	0.045
	Solid to gas mass ratio entering to cyclone, C_e [-]	3.8
	Gas velocity at the inlet section of the cyclone, u_e [m/s]	14.83
	Cyclone dimensions [m]: $Dc = 0.182$; $A = 0.080$; $C = 0.038$; $M = 0.073$; $F = 0.091$; $S = 0.255$; $B = 0.709$; $N = 0.073$.	-
	Gas velocity at cyclone exit, v_o [m/s]	10.90
	Pressure drop in the cyclone, ΔP_c [Pa]	271
Standpipe and L-valve	Internal diameter of the standpipe and the L-valve, D_{lv} [m][(1)]	0.063
	Circulated solids flux in the standpipe and the L-valve, $G_{s,lv}$ [kg/m^2.s]	55.2
	Solids velocity in the moving bed, $U_{s,lv}$ [m/s]	0.04
	Horizontal section of the L-valve $L_{h,lv}$ [m]	0.36
	Aeration level above of the horizontal section center line, L_a [m]	0.13
	Aeration mass flow rate in the L-valve, m_{lv} [kg/h]	1.97
	Height of the solids above of the aeration point, $L_{v,lv}$ [m]	0.50
	Height of the standpipe, L_{sp} [m]	2.71
	Pressure drop in the L-valve, ΔP_{lv} [Pa]	3,917
	Pressure drop in the standpipe, ΔP_{sp} [Pa]	7,265
	Total solids inventory in the CFB system [kg]	8.20

[(1)] The value of the actual system was adopted for the solids inventory comparison.

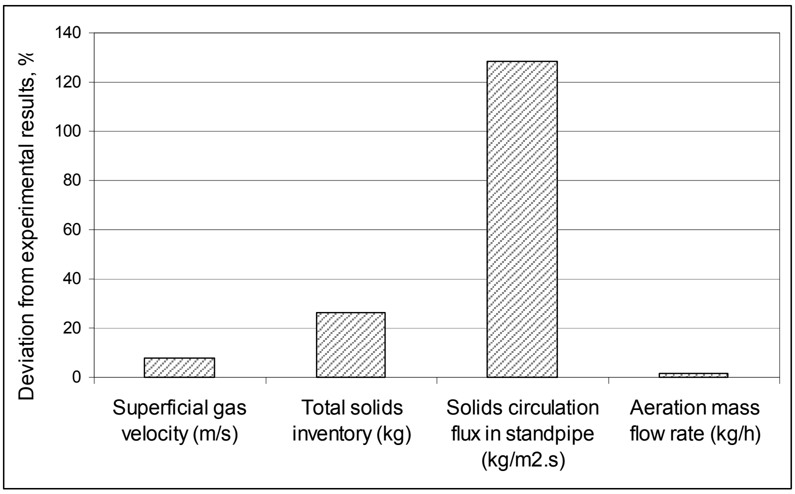

Figure 3. Validation of the semi-empirical model for the main CFB operational variables.
Comparison from Hory, *et al*. (2006).

The Fig. 3 shows that the model predicts the experimental conditions within acceptable deviation for U, $I_{s,T}$ and m_{lv}. The better approximation and the worse one were found for the aeration mass flow rate and the solids circulation flux, respectively. For the total solids inventory, results obtained in the simulation were overestimated in approximately 26%. In general, these results are considered satisfactory for preliminary design of CFB loops.

4. CONCLUSIONS

In this work, the structure of a practical semi-empirical model for bench-scale CFB systems based on hydrodynamics relationships and experimental data from literature was proposed. The results obtained from a design case and validation under pre-heating condition using quartz sand as the bed material suggest that the model predicts the main dimensions and configuration of the CFB loop satisfactorily, as well as, the basic operational parameters needed for stability at high temperature and atmospheric pressure.

The proposed semi-empirical model can be easily implemented and used as useful tool for the preliminary design and pre-heating operational setup of bench-scale biomass CFB combustors or gasifiers. The model validation at higher scales needs to be verified looking forward commercial application.

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