Dynamic modeling of the circulating fluidized bed riser

Rupen Panday^{a,b}, Ronald Breault^{a, 1}, and Lawrence J. Shadle^a,

^aNational Energy Technology Laboratory

^bREM Engineering Services PLLC

3610 Collins Ferry Road, Morgantown, WV-26507, USA

ABSTRACT

Dynamic modeling is required to understand hydrodynamics of a circulating fluidized bed (CFB) riser as a function of time. The paper develops a dynamic model of the riser from the conservation of solids mass. Solids mass that flows out of the riser requires particle residence time to close the mass balance equation. The mean residence time of solids was derived as the riser length divided by the superficial riser gas velocity normalized by a ratio of solids concentration in the riser outlet to that in the riser itself. For bed materials including high-density polyethylene beads, glass beads, and cork, the percentage of solids concentration in the horizontal crossover pipe connecting the riser exit and the cyclone inlet was found to be 22% of the average concentration of solids in the riser. Upon finding the particle residence time, we calculated the solids flow rate out of the riser and along with the knowledge of solids flowing into the riser column we determined the riser inventory as a function of time. We used the estimated solids inventory to model overall pressure drop across the riser from the conservation of momentum. The momentum equation included the balance of pressure force by the hydrostatic head of solids, and the wall frictions due to gas and solids in a fully developed zone. The prediction of overall riser pressure drop compared well with its measured values at many dynamic conditions. Therefore, our dynamic model is useful for developing advanced control strategy or for improving existing control systems of the CFB plant. The residence time approach is also useful for the design of a transport reactor.

¹ Corresponding Author, Tel.: +1 304 285 4486.
Email Address: ronald.breault@netl.doe.gov

The technique also comes handy when setting up the boundary condition for the overall pressure drop in the CFD modeling that can simulate incremental pressures along the riser height.

Keywords: Circulating Fluidized Bed, Inherent Feedback, Riser State Model, Average Residence Time, Dynamic Experiments

1. Introduction

Circulating fluidized beds have many industrial applications including coal gasification, oil refining, Fischer-Tropsch synthesis, and several other chemical processes that use solids circulation system and employ one or more fluidized beds. A circulating bed can transport large quantities of material thereby transferring a significant amount of heat required for different applications. A circulating fluidized bed (CFB) consists of a riser and a standpipe in a closed loop. The two components are arranged vertically beside one another (Figure 1). At the bottom of the loop, the solids flow from the standpipe, also called a downcomer, into the riser. In Figure 1, the solids transfer is motivated by means of aerating the non-mechanical L-valve. In other installations, a mechanical valve such as a slide valve is used. The gas supplied at the riser bottom transports the solids to its top. After exiting the top, these solids are directed horizontally by a crossover pipe and injected tangentially into a cyclone. The cyclone separates the solids from the gas that exits through its top, and the solids fall back into the downcomer. The cycle repeats thus creating a continuous circulation of solids around the CFB loop.

In the past, many researchers have attempted to characterize the hydrodynamics of the circulating fluidized bed (CFB) risers. Even the competitions involving hydrodynamic modeling of the CFB riser were arranged in the form of challenge problems for the last 20 years [1], [2], [3]. Researchers working in the multiphase flow area participated in these challenges. The strengths and weaknesses of their models were investigated against the experimental data collected during steady states. However, there have not been many developments of dynamic models describing the riser hydrodynamics compared to its steady

state counterparts. Dynamic models help understand the riser hydrodynamics as a function of time that eventually helps developing better control strategy and CFB optimizations.

In circulating fluidized bed systems, reactions are normally conducted in the riser and for catalytic processes, the catalyst regeneration is performed in the vessel fed by the downcomer and recycled back to the riser. In this respect, the solids circulation rate, the superficial riser gas velocity and the system inventory are the key operating variables. They determine the pressure drop across the riser as well as across other components of the circulating system. For calculating mass inventory in the riser, void fraction or voidage is needed. The fraction of column cross-sectional area available for flow of gas is determined by the volume fraction occupied by gas, ε_r . As a result, the fraction of pipe area available for the flow of solids is $(1 - \varepsilon_r)$. Experimentally, the total solids inventory in the riser can be determined in several ways. In one way, the CFB unit is brought to a steady state at a given operating condition. Then, the solid inventory is obtained by shutting off the gas and solids feed into the riser, and draining and weighing the accumulated solids in the riser. In another way, the solids feed to the riser is cut off by shutting off the standpipe and L-valve aeration flows. The riser column is allowed to empty by continuously feeding the gas at its bottom. The amount of solids leaving the riser top accumulates in the standpipe. The difference between the bed heights in the standpipe before and after solids cut off provides the mass of solids initially present in the riser. Both of these techniques require shutting down either the solids flow in the CFB or the gas flow in the riser. As a result, the solids circulation ceases around the loop. Therefore, these techniques are impractical in critical applications such as establishing riser inventories in the utility power distribution. Coal fired power plants burn coal in the CFB boiler for transferring heat to the water boiler drum, which turns the steam turbine coupled to the electrical generator. The amount of coal to be burnt in the CFB is determined by the customer demand. During a high demand period, more coal is needed and hence, it is impractical to shut down the CFB boiler to meet the customer demand.

There are studies that have attempted to continuously monitor riser inventory [4, 5, 6, 7, 8] during the operation of CFB units. The common technique involves determining solids inventory from pressure drop measurements. The solid inventory is calculated by assuming that the weight of solids balances the riser pressure drop. The shear stress at the wall is usually neglected even though it may contribute significantly to the total pressure drop [9]. Most of these analyses focused only on the steady state conditions in which, by definition, the rate of change of riser inventory is zero. As a result, the mass flowing into the riser equals the mass flowing out.

During transients, the mass balance equation is described by a partial differential equation rather than an algebraic expression. In the present paper, the cross-sectional averaged values of the process variables were considered. Therefore, the mass balance model is described by the ordinary differential equation. We took the entire riser of a cold flow CFB system as the control volume for our analysis. In cold flow setup, the chemical reactions do not take place and, hence, at a given superficial gas velocity, U_g , the mass balance equation for the riser inventory becomes:

$$\frac{dM_r(t)}{dt} = M_{s,in}(t) - M_{s,out}(t) \quad (1)$$

where $M_{s,in}(t)$ and $M_{s,out}(t)$ represent the mass flowing into and out of the riser, respectively at any time t . The left hand side is the time rate of change of solids ($M_r(t)$) within the riser.

At steady state, the left hand side of Eq. 1 becomes zero.

$$\frac{dM_r(t)}{dt} = 0 \quad (2)$$

Both mass flow rates equal the steady state solids circulation rate (M_s). That is,

$$M_{s,in}(t) = M_{s,out}(t) = M_s \quad (3)$$

The riser inventory would be maintained at some fixed value, M_r or,

$$M_r(t) = M_r \quad (4)$$

When solids enter the riser, they spend some time in the riser volume before leaving the top along with the gas. The average amount of time that solids spend in the column is defined as the average residence time (τ_r) of the solids. It is given by

$$\tau_r = \frac{M_r}{M_s} \quad (5)$$

If it is assumed that $M_{s,in}$ is known and τ_r is fixed at a given condition, then, from Eqs. 3 and 5, $M_{s,out}$ becomes:

$$M_{s,out}(t) = \frac{1}{\tau_r} M_r(t) \quad (6)$$

Substituting Eq. (6) in Eq. (1) gives

$$\frac{dM_r(t)}{dt} = M_{s,in}(t) - \frac{1}{\tau_r} M_r(t) \quad (7)$$

Eq. 7 is the dynamic model of the riser at a given superficial gas velocity. Under normal operating conditions, the mass flow rate into the riser can be varied and the time variations in the riser inventory can be observed. During transients, the mass flow rate varies with time and hence, mass inventory in the riser will change with time. The time-dependent inventory, $M_r(t)$, is actually the state variable. In other words, not only does it require knowledge of the solids inflow rate and the solids residence time, but also knowledge about the prior riser inventory.

In terms of velocity, the mean residence time (τ_r) of solids is defined as:

$$\tau_r = \frac{\Delta h_r}{u_p} \quad (8)$$

where Δh_r is the length of the riser and u_p is the net velocity of particles flowing upwards with respect to the riser wall such that eventually all the particles will be transported out of the riser. The first choice for u_p would be to calculate its value from the slip velocity (u_{slip}) that represents the interaction between gas and particles arising from the drag force on the particles. Kunii and Levenspiel [10], W. C. Yang [11] have used the slip velocity as a parameter to model the hydrodynamics of experimental risers.

$$u_{slip} = u_g - u_p = \frac{U_g}{\varepsilon_r} - u_p \quad (9)$$

where ε_r is the average voidage across the riser cross-sectional area (A_{rs}) and u_g is the interstitial gas velocity. In Eq. 9, the frame of reference is the riser wall and if the upward direction is chosen positive, both interstitial gas velocity (u_g) and average particle velocity (u_p) are positive as they travel upwards with respect to the riser wall. The value of u_{slip} is assumed to be equal to the particle terminal velocity (u_t) from which the particle velocity is calculated [12]. The assumption is valid only in a dilute regime where the influence of the nearby solids on a particle may be neglected when compared to the gas-particle interaction [13], [14]. Patience et. al. [15] also showed that the assumption, $u_{slip} = u_t$, is not true and concluded that better criteria are required to evaluate the average solids velocity. We used a criterion on choosing a flow condition as a dilute flow based on the influence of solid particles on the drag force exerted by the gas. Gidaspow [13], Yang [11] suggested calculating the drag coefficient for a single particle first and then to multiply the calculated drag coefficient with a factor like $\varepsilon_r^{-2.65}$. That is, if the amount of solids in the riser is 10% i.e., $\varepsilon_r = 0.90$, then the single particle drag coefficient (C_d) should be modified as $C_d \cdot 0.90^{-2.65} = 1.32 C_d$. Thus, the 32% correction on C_d accounts for the influence of nearby particles on the drag force. In dilute flow, the voidage is closer to 1, say $\varepsilon_r = 0.98$, that makes the correction $\varepsilon_r^{-2.65}$ closer to 1. Therefore, the drag term $C_d \cdot 0.98^{-2.65}$ becomes closer to a single particle drag coefficient, C_d and a single particle terminal velocity (u_t) may be substituted for the slip velocity, u_{slip} in dilute flow conditions.

Patience et. al. [16] defined a slip ratio (ψ) as the ration of the interstitial gas velocity (u_g) to the average particle velocity (u_p).

$$\psi = \frac{u_g}{u_p} = \frac{U_g}{\varepsilon_r u_p} = \frac{U_g \rho_p (1 - \varepsilon_r) A_{rs}}{\varepsilon_r M_s} \quad (10)$$

where, ρ_p is the particle density. Eq. 10 has two unknowns - ψ and ε_r - for the given values of solids circulation rate (M_s), superficial gas velocity (U_g), particle density and the cross-sectional area of the riser (A_{rs}). To calculate slip ratio (ψ), Patience et. al. proposed a following correlation (Eq. 11) as a function of

Froude Number (F_{rt}) based on a particle terminal velocity (u_t) and that of a Froude Number (F_r) consisting of a superficial gas velocity.

$$\psi = 1.0 + \frac{5.6}{F_r} + 0.47F_{rt}^{0.41}, F_r = \frac{U_g}{\sqrt{gD_r}}, F_{rt} = \frac{u_t}{\sqrt{gD_r}} \quad (11)$$

where g is the acceleration due to gravity and D_r is the internal diameter of the riser. They suggested calculating the slip ratio from Eq. (11) first, and, using it to get the void fraction (ϵ_r) from Eq. (10). Then, the resulting voidage (ϵ_r) gives the average particle velocity (u_p) from the solids circulation rate (M_s).

$$u_p = \frac{M_s}{\rho_p(1-\epsilon_r)A_{rs}} \quad (12)$$

The slip factor was calculated from Eq. 9 using the particle terminal velocity calculated from the correlation given by Kunii and Levenspiel [10]. The residence time (τ_r) was determined by substituting the value of u_p in Eq. 11 and the total solids inventory (M_r) within the riser was found by using the calculated value of τ_r in Eq. 5. Then, the calculated solids mass (M_r) was used to determine the overall pressure drop across the riser (ΔP_r) considering only hydrostatic head of solids and neglecting wall friction contributions due to gas and solids [16].

$$\Delta P_r = \frac{M_r g}{A_{rs}} \quad (13)$$

The calculated pressure drop from Eq. 13 was compared against the experimental data collected in the circulating fluidized bed present at the National Energy Technology Laboratory (NETL) (Figure 1).

Figure 2 shows the comparison of the measured and calculated pressure drops for high-density polyethylene beads (HDPE), glass beads (GB) and cork particles, circulated at a wide range of solids flow rates and superficial gas velocities. The experiments were conducted at $0.3 \leq M_s^{HDPE} \text{ (kg/s)} \leq 18.8$, $7.6 \leq U_g^{HDPE} \text{ (m/s)} \leq 14.0$; $2.7 \leq M_s^{GB} \text{ (kg/s)} \leq 22.6$, $6.2 \leq U_g^{GB} \text{ (m/s)} \leq 10.8$; $0.2 \leq M_s^{Cork} \text{ (kg/s)} \leq 1.3$, $4.1 \leq U_g^{Cork} \text{ (m/s)} \leq 10.9$. The model obtained from Eqs. 10-13 predicted only 50% of the experimental value at the variance (R^2) of 96%, 89% and 98% for HDPE, GB and cork respectively. In summary, Patience

approach underestimates the average residence time of solids (Eq. 8) thus predicting a lower pressure drop across the riser.

Patience et. al. developed a correlation (Eq. 11) in the hydrodynamically fully developed region of the circulating fluidized bed riser and neglecting the acceleration zone at the bottom and the deceleration region at the top of the risers. The assumption is not viable because, depending on the superficial riser gas velocities, all three zones exist simultaneously in a 15.4 m tall NETL riser that is equipped with a blind Tee at its top. The fully developed region lies above the acceleration zone from the riser entrance and below the deceleration zone near the riser exit. The acceleration and deceleration regions are characterized by a denser region whereas the bed density is leaner in the middle developed section. When solids enter the riser, they are accelerated up to a certain elevation due to high gas flow rate given at the riser base. Beyond the acceleration length, the void fraction profile becomes nearly constant. When the riser exit is at a sharp angle to the upward flowing gas and solids, the top of the riser acts as a crude gas-solids separator [17]. The particles cannot negotiate the sharp turn and reflect off the top. They re-circulate internally down along the walls of the column. Consequently, an area of densification is created around the exit. Depending on the solids circulation rate and severity of the exit restriction, the deceleration length may extend well below the riser top. Figure 3 shows apparent void fraction profile in the column for a superficial gas velocity of 13.72 m/s and various solids circulation rates. The apparent voidage was calculated from the pressure drop measurements made at 20 axial locations along the riser height. These are apparent voidages, that is, only the contribution of hydrostatic heads of the solids was considered. In Figure 3, green diamond and light blue dash symbols represent small or no entrance and exit effects arising from the blind Tee at the riser top. Orange triangle and brown plus signs show moderate entrance effect and densification of solids taking place around the exit. Blue circle and red cross marks show more profound entrance and exit effects. They represent an extreme case where the entrance and exit effects propagate throughout the riser length.

The experimental data of Figure 3 reveals that the riser entrance and exits have significant effects on gas-solids flow characteristics. Therefore, the residence time should be representative of the entire riser, not only of its fully developed zone. However, modeling detailed hydrodynamics of the entire riser is a difficult task that requires not only the knowledge of acceleration and deceleration lengths but also involves complex movements of the solids particles in the non-uniform flow fields. Many authors (Weinstein and Lee [18], Monazam and Shadle [19], Zenz and Weil [20]) described the voidage profile in the acceleration zone as an exponentially decaying solids fraction. To account for the deviation in the top solids fraction arising from an abrupt exit, the solids fraction at the riser top is determined assuming a smooth exit [21] first, and, then, corrected by an exponential expression [22]. The exponential correction requires the knowledge about decay constant and reflux constant at the riser top that can be calculated from the empirical correlations given in Kim et. al. [23].

2. Solids Residence Time in the Riser

Generally, the residence time distribution of gas molecules in chemical reactors like plug flow reactor, continuously stirred tank reactor and others is determined experimentally by injecting a tracer at the inlet and then measuring the tracer concentration in the effluent stream as a function of time. The mean residence time is subsequently calculated as the first moment of the residence time distribution. In the present work that involves gas-solids interaction in the riser, the average residence time of solids was calculated considering the riser outlet condition. First, we assumed a horizontal crossover pipe connects the riser exit at the top with the cyclone inlet. When gas and solids exit the vertical riser, the particles accelerate along the horizontal pipe along with the gas. The solids concentration is usually very low in the crossover [24], [25], [26], [27]. With a dilute flow of solids existing in the crossover, it is assumed that the average particle velocity ($u_{p,co}$) attains the value of the superficial gas velocity ($U_{g,co}$) suggesting a zero slip between the gas and solids (Eq. 9) in the crossover. Then, the residence time (τ_r) defined in Eq. 5 can be manipulated as:

$$\tau_r = \frac{M_r}{M_{s,out}} = \frac{\rho_p(1-\varepsilon_r)A_{rs}\Delta h_r}{\rho_p(1-\varepsilon_{co})A_{co}u_{p,co}} = \frac{(1-\varepsilon_r)}{(1-\varepsilon_{co})} \frac{A_{rs}}{A_{co}} \frac{\Delta h_r}{U_{g,co}} \quad (14)$$

where, ε_{co} is the average voidage in the crossover pipe and A_{co} is the cross-sectional area of the crossover. The superficial riser gas velocity (U_g) is computed from the volumetric flow rate of gas given at the riser bottom and those coming from the standpipe side. The gas velocity (U_g) is usually corrected for the riser operating pressure and temperature, but the pressure drop across the crossover is small compared to that in the riser for a given operating condition. Assuming there are no volumetric gas flow rate in the crossover other than the one coming from the riser, the superficial gas velocity in the crossover ($U_{g,co}$) is chosen to be the superficial gas velocity in the riser (U_g) multiplied by a ratio A_{rs}/A_{co} . As a result, Eq. 14 becomes

$$\tau_r = \frac{1}{\frac{(1-\varepsilon_{co})}{(1-\varepsilon_r)}} \frac{\Delta h_r}{U_g} = \frac{1}{\kappa} \frac{\Delta h_r}{U_g} \quad (15)$$

where, the ratio, $\kappa = I-\varepsilon_{co} / I-\varepsilon_r$, represents the average concentration of solids in the crossover with respect to the average solids concentration in the riser.

In Eq. 15, the parameter κ is unknown. The value of κ was calibrated from the riser inventory (M_r), solids circulation rate (M_s), and superficial riser gas velocity (U_g) at steady state conditions. In Figure 4, the diamond represents the HDPE particles circulated at the solids circulation rate between the range of 0.3 kg/s and 18.8 kg/s and the superficial gas velocity ranging from 7.6 m/s to 14.0 m/s. The circle represents glass beads operated with the solids circulation rate in the range of 2.7 kg/s and 22.6 kg/s at the superficial gas velocity maintained between 6.2 m/s and 10.8 m/s. The triangle indicates cork particles operated between 0.2 kg/s and 1.3 kg/s of the solids circulation rate at the superficial gas velocity varying from 4.1 m/s to 10.9 m/s. The value of a parameter κ was calibrated between 19% and 23% depending on the bed material. A variance (R^2) was explained between 97% and 99% of the experimental uncertainty (Figure 4a). Figure 4b consists of the data sets of these three bed materials into a single plot where a common value of 22% was attained for a parameter κ at R^2 of 98%.

To estimate the value of κ , the riser inventory (M_r) is required. The solids inventory was calculated from the total pressure drop (ΔP_r) measured across the riser. The pressure drop was modeled based on the momentum balance equation that is given below:

$$\Delta P_r = \frac{g}{A_r} M_r + 2f_s \frac{\rho_p(1-\varepsilon_{fd})u_{pfd}^2 \Delta h_r}{D_r} + 2f_g \frac{\rho_g U_g^2 \Delta h_r}{\varepsilon_{fd} D_r} \quad (16)$$

In Eq. 16, the contributions of acceleration and deceleration of gas and solid particles were not considered. The static head due to gas was neglected because it is small in comparison with the solids contribution. The voidage (ε_{fd}) and the particle velocity (u_{pfd}) appearing in the friction contributions of Eq. 16 were their average values in the fully developed region of the riser. When particles reach the hydrodynamically developed zone, their velocities (u_{pfd}) remain relatively fixed. Then, the balance of momenta for solids alone gives ([13], [14])

Drag Force = Gravitational Force + Wall Friction due to Solids

$$\text{or, } \frac{3}{4} \frac{C_d \varepsilon_{fd}^{-2.65}}{d_p} \frac{\rho_g}{(\rho_p - \rho_g)} \left| \frac{U_g}{\varepsilon_{fd}} - u_{pfd} \right| \left(\frac{U_g}{\varepsilon_{fd}} - u_{pfd} \right) = g + 2f_s \frac{u_{pfd}^2}{D_r}, \varepsilon_{fd} \geq 0.8 \quad (17)$$

where, C_d is the drag coefficient, d_p is the sauter mean particle diameter, ρ_g is the density of gas and f_s is the Fanning friction factor for solids. The change in gas density due to change in pressure in cold flow experiments was negligible compared to the magnitude of particle density. Hence, the standard value of ρ_g ($= 1.2 \text{ kg/m}^3$) was used throughout the analysis. The particle velocity (u_{pfd}) can be obtained from Eq. 17 provided the values of C_d , U_g , ε_{fd} , f_s , d_p , and D_r are known. Except the values of C_d , f_s and ε_{fd} , measured values of U_g , d_p and D_r were available. The drag coefficient (C_d) was computed from the following correlation [13],

$$C_d = \begin{cases} \frac{24}{Re_s} (1 + 0.15 Re_s^{0.687}), & Re_s < 1000 \\ 0.44, & Re_s \geq 1000 \end{cases} \quad (18)$$

where, the Reynolds Number (Re_s) was

$$Re_s = \frac{\varepsilon_{fd} \rho_g \left| \frac{u_g}{\varepsilon_{fd}} - u_{pfd} \right| d_p}{\mu_g} \quad (19)$$

The Fanning friction factor (f_s) was calculated from Konno and Saito's relation [28]

$$f_s = 0.0285 \frac{\sqrt{g D_r}}{u_{pfd}} \quad (20)$$

and the void fraction (ε_{fd}) was determined from the solids circulation rate, i.e.,

$$\varepsilon_{fd} = 1 - \frac{M_s}{(\rho_p A_{rs} u_{pfd})} \quad (21)$$

Eq. 17 is nonlinear in u_{pfd} and a generalized reduced gradient solver was used to solve for u_{pfd} .

In Eq. 16, the wall friction contributions to the overall riser pressure drop (ΔP_r) were assumed same as those in the fully developed zone. The solids friction factor was calculated from Eq. 20. The gas friction factor was determined from the standard Stokes law, Blasius relation and transmission formula at different Reynolds number[13]. In calculating gas friction coefficient, Reynolds number used the riser diameter, rather than the particle diameter (Eq. (19)), as the characteristic length. Thus,

$$f_g = \begin{cases} \frac{16}{Re_g}, & Re_g \leq 2100 \\ \frac{0.0791}{Re_g^{0.25}}, & 2100 < Re_g \leq 100,000 \\ \frac{1}{\left(2 \log \left(Re_g \sqrt{f_g} \right) \right)^{0.8}}, & Re_g > 100,000 \end{cases} \quad (22)$$

where the Reynolds number (Re_g) was

$$Re_g = \frac{\rho_g U_g D_r}{\mu_g} \quad (23)$$

for a given gas density (ρ_g) and gas viscosity (μ_g).

3. Results and Discussions

3.1 Experimental Facility

The state model (Eq. 7) developed in this work was compared against the experimental data collected at different dynamic conditions. The dynamic experiments were conducted in the cold flow circulating fluidized bed system shown in Figure 1. The cold flow CFB is an industrial scale unit that is mainly used for hydrodynamic studies of gas-solids flow. The unit is also used for CFD modeling of gas-solids flow hydrodynamics and the model validation of the transport gasifier.

The CFB unit consists of a 15.4 m tall riser with an internal diameter of 0.305 m. The standpipe is 13 m tall and has a 0.254 m inside diameter. Transport air is introduced at the bottom of the riser through a perforated plate distributor. The distributor section is located 0.43 m below the centerline of the L-valve. There are eight aeration ports at different elevations in the standpipe. The L-valve is equipped with a sparger. The sparger gas lifts the solids when a bed of solids moves from the standpipe to the riser. The solids circulation measurement device, known as the Spiral, is located approximately, 3.5 m above the entrance of the horizontal L-valve. The spiral rotates as the solids flow down through it and generates a continuous measurement of average solids velocity in the CFB standpipe [29], [30]. The area of the standpipe multiplied by the solids velocity provides volumetric solids flow measurements. The volumetric solids flow was converted into a mass circulation using the density of a moving bed. The bed density was calculated from the standpipe model [31], [32] that uses the pressure drop measured across the spiral location. The standpipe model predicts the void fraction along the standpipe height as well as the magnitude of gas flowing up the standpipe and/or down into the riser. The latter flow, and the L-valve sparger, were combined with the riser gas flow to calculate the superficial velocity (U_g) at the standard condition (101 kPa & 20 °C). Riser gas velocity was then corrected for temperature and pressure measured at the base of the riser. There are many pressure drop measurements throughout the circulating loop. During experiments, the relative humidity of air was maintained between 40 and 60%. Occasionally, Larostat was added in the riser. This antistatic agent along with the humidity helped minimize the buildup of electrostatic charges in the unit due to solids rubbing against the wall. Most of the riser and standpipe sections are made up of clear acrylic plastic. The clear sections allow advanced measurements such as

high-speed imaging of particles [33] moving in the gas stream. During operation, a steady state condition of the circulating fluidized bed was defined as maintaining a constant set of flow conditions and a constant response in the riser pressure drop over a five-minute period. The data was acquired every second for all the variables except the standpipe bed height. Bed height was visually recorded at the end of each steady state condition. When bed level lied out of view in the steel section, it was calculated from the standpipe pressure measurements [31].

3.2 Bed Materials

Three bed materials including HDPE beads, glass beads, and cork were used to generate the transient data in the circulating fluid bed. The CFB unit had L-valve while conducting experiments with the HDPE particles and glass beads. The cork data set were collected using a loopseal as the non-mechanical valve between the CFB riser and the standpipe (not shown in Figure 1). HDPE beads are the Group B materials according to Geldart's classification [34]. These polyethylene beads were coated with a phosphorescence dye to enable optical experiments to be conducted on moving particles. In these experiments, light is flashed on the moving solids at a specific location. The phosphorescent coating absorbs and re-emits the incident light. The re-emitted light intensity on the moving particles are then measured by the photomultipliers installed at specific location in the CFB unit. Because of the phosphorescent coating, HDPE beads have wide particle size distribution [35]. Another bed material namely glass beads are heavier and significantly smaller than HDPE beads, and belong to Geldart Group A classification. The last bed material, cork, is the lightest particle among the three and falls into Geldart Group B category. The properties of HDPE, glass beads and cork are given in Table 1. The nominal void fraction (ε_n) listed in Table 1 represents a value intermediate between the fully packed granular bed (ε_p) and a fluffed bed (ε_f). The nominal value corresponded to a voidage determined from a benchtop experiment. The nominal voidage was initially used for solids circulation rate calculation that was later replaced by the voidage obtained from the standpipe model [31]. The standpipe model uses pressure drop measurements in the standpipe during operation. Hence, the voidage from the standpipe model may better

represent the bulk density of solids than the nominal voidage. The nominal values were measured using a standard pycnometer with water as the displacement fluid [36]. The particle size was estimated by a sieve analysis using the ASTM C-136-06 Standard. The minimum fluidization velocity was determined in a 0.05 m diameter cylindrical column. The values of the transition velocities (U_{tr1} and U_{tr2}) were calculated from the data collected in the circulating fluidized bed riser [37] used in this study. The lower transport velocity (U_{tr1}) corresponds to a classical choking velocity while the upper transport velocity (U_{tr2}) separates the homogeneous transport condition from the fast-fluidized regime. In Table 1, the dimensionless quantities including Archimedes Number (Ar), Reynolds (Re_t) and Froude Numbers (Fr_t) based on particle terminal velocity, and the ration of particle diameter to the riser diameter commonly used in fluidized bed scale up [38] are also presented.

3.3 Dynamic Data Analysis - Inherent Feedback Mechanism in the Circulating Fluidized Bed

Circulating fluidized beds are inherently the closed loop systems. Cold model under consideration is governed by both mass and force balances across various components of the loop. For operation, control, and stability analysis of the CFBs, it is necessary to understand the hydrodynamic coupling between the riser and the standpipe.

The pressure balance in the circulating fluidized bed unit with an L-valve is given by the following equation:

$$\Delta P_{sp} = \Delta P_{lv} + \Delta P_r + \Delta P_{co} + \Delta P_{cy} \quad (24)$$

$$\Delta P_{sp} \approx \Delta P_{lv} + \Delta P_r \quad (25)$$

The pressure drop across the crossover (ΔP_{co}) and the cyclone (ΔP_{cy}) are small compared to the other individual pressure drop and therefore, Eq. 24 reduces to Eq. 25. Eq. 25 suggests that a CFB system always possesses a feedback relation between the riser and the standpipe and vice versa. A change in

pressure drop across the standpipe (ΔP_{sp}) affects the pressure drop across the L-valve (ΔP_{lv}) and the riser (ΔP_r). Similarly, a change in the riser pressure drop affects the pressure drop across the standpipe.

There is a maximum limit of pressure drop that the standpipe can develop. In the standpipe, solids are transferred by gravity from a low pressure to a higher pressure. The aeration gas in the standpipe flows upward relative to the downward moving solids in the standpipe. The relative gas-solids flow will then generate the sealing pressure drop required for maintaining stability of the overall system. The pressure drop across the standpipe increases with the relative velocity between the gas and solids. The maximum pressure drop that the standpipe can generate is equal to the pressure drop at its incipient fluidized state, $(\Delta P_{sp}/H_{sp})_{mf}$. When the relative velocity becomes higher than the minimum fluidization velocity (U_{mf}), bubbles are generated. Any bubble formation, in turn, may hinder the particle flow and reduce the solids flow rate. These bubbles, therefore, lead to unstable feeding of solids from the standpipe into the riser.

The riser pressure drop, ΔP_r , increases with an increase in the solids circulation rate at a given superficial gas velocity [39]. If a small reduction in gas velocity takes place for a given solids circulation rate, the riser pressure drop increases. According to Eq. 25, this increase in ΔP_r has to be balanced by the pressure build up in the circulating loop. If the relative velocity between gas and solids are below the minimum fluidization velocity of the solids, the pressure drop across the standpipe can be adjusted accordingly. The operating state of the CFB will then be shifted to a new balance point. However, if the riser and L-valve pressure drops become higher than the maximum pressure drop available from the downcomer, the system will start to oscillate [40]. The transport gas in the riser flows into the standpipe. Such flow reversal must be avoided at all times.

During operation, the initial system inventory (M_i) splits into different components of the circulating fluidized bed. The overall mass balance equation is given by:

$$M_t = M_{sp} + M_{lv} + M_r + M_{co} + M_{cy} \quad (26)$$

The solids inventory in the crossover (M_{co}) and the cyclone (M_{cy}) are neglected because they are in transit or because they are small in quantity compared to the amount of solids present in the riser or in the standpipe. The inventory in the standpipe (M_{sp}), riser (M_r) and the L-valve (M_{lv}) are of primary interest in the CFB analysis. Therefore, Eq. 26 is re-written as:

$$M_t \approx M_{sp} + M_{lv} + M_r \quad (27)$$

At any instant, the height of moving solids in the standpipe sets the standpipe inventory. Both the standpipe bed height and the riser gas velocity determine the limit of solids circulation rate in the system. In order to achieve desired solids circulation rate at a given riser gas velocity, the standpipe bed height cannot be dropped to a limit that leads to the standpipe condition where the standpipe cannot develop sufficient pressure head to push the solids into the riser. If the pressure build in the standpipe is not high enough, the standpipe cannot transfer solids at a rate high enough to equal the saturation carrying capacity of the riser gas [41]. Hence, the riser empties out quickly and the solids recirculation ceases. The higher standpipe bed level alone, however, cannot guarantee the desired solids circulation rate. If the riser gas velocity is kept low, a dense bed will be created at its bottom [42]. The duration to form the dense bed when the gas velocity is lowered depends on the prior inventory residing in the riser. On keeping the superficial velocity further below the lower transport velocity (U_{trl}), a plugging of riser due to solids starts taking place at or near the solids entry point in the riser. The transport gas cannot take enough solids out of the riser at a rate that is equal to the rate of solids coming into the riser from the standpipe. The gas pressure in the riser increases which results in the appearance of solids slugs right after a dense bed is created in the riser bottom. The slug flow causes the whole system to fluctuate dramatically and stable operation is no longer possible. Slugging in the riser is one mechanism whereby the dependent variable, such as the riser pressure drop, leads the solids flowing into the riser.

At NETL CFB, the bed level was always kept above the spiral location in order to avoid exposure of the device. In this work, the low bed height referred to the bed level above the spiral location but it

does not correspond to the limit where the standpipe cannot develop sufficient pressure head as discussed above. The low bed height generally referred to 5.2 m above the entrance of horizontal L-valve and the high bed level corresponded to a level at or above 9.2 m from the standpipe bottom. When the standpipe was operated at low bed height, the dependent variable such as solids flow rate measured in the standpipe led the aerations that actually moved solids from the standpipe to the riser [43]. The phenomenon can be ascertained by referring to Figure 5. The data were collected under low and high system inventories when the riser was operated in homogeneous transport regime. If the superficial gas velocity is kept above the upper transport velocity (U_{tr2}), the distribution of solids becomes homogeneous in the riser [37]; hence the name homogeneous transport regime. The superficial gas velocity between U_{tr1} and U_{tr2} gives the operating regime that is commonly known as fast-fluidized bed. The fast-fluidized regime is characterized by the S-shaped voidage profile along the riser height [44], [45]. On the other hand, the homogeneous transport regime typically shows a C-shaped voidage profile when the riser is equipped with an abrupt exit (Figure 2).

In Figures 5a and 5c, the plots included the filtered time series data of the solids circulation rate (black line) and the aeration flow (green line). The time series data of the calculated bed height (blue line) and aerations (green line) are presented in Figures 5b and 5d. Here, the standpipe aerations were modulated at 60 seconds period and both the solids circulation rate and the bed height showed the oscillations at 60 seconds period. The fluctuations in the spiral data are usually high and 5 periods that correspond to the total duration of 5 minutes were averaged together to generate the filtered time series of circulation rate in Figures 5a and 5c. The bed height was calculated from the standpipe pressure measurements using a technique proposed by Ludlow et. al. [31]. The bed height data have less fluctuation compared to the solids circulation rate measurements but has a phase difference of 180 degrees with respect to the sinusoidal aerations. Higher aerations mean lower standpipe bed height because more solids start moving towards the riser. At less aerations, solids start to build up in the standpipe due to solids coming in from the riser thus giving higher bed height. Therefore, the bed height

signal was completely out of phase from the aerations as can be seen from Figures 5c and 5d. In Figures 5a and 5b, the plots refer to the low system inventory corresponding to the average standpipe bed height of 5.2 m, high superficial gas velocity (10.7 m/s) and high solids circulation rate (5.9 ± 0.6^2 kg/s). The plots in Figure 5c and 5d denote the same riser operating condition as the first one. However, the high system inventory was used giving the standpipe bed height of 10.0 m and the solids circulation rate of 6.8 ± 0.7 kg/s. At high bed height, there was no phase difference between the solids circulation rate and the standpipe aeration on the average (Figure 5c). The solids circulation rate led the standpipe aeration by 3 seconds in the inventory limiting case (Figure 5a). The phase angles were calculated from the maximum of Fast Fourier Transform (FFT) of both the standpipe aeration and solids flow rate signals. The time difference (Δt) was calculated from the difference in phase angles between these two variables.

In CFB operation, it is the aeration that initiates solids movement from the standpipe. The aerations' effectiveness in moving solids depends on the standpipe bed height. At high standpipe bed height (10.7 m), the aeration effectively produced a given solids circulation rate and, higher the solids level in the standpipe, lesser the aeration needed to generate a given solids flow rate. The fact is illustrated in Figure 5c where less aeration ($F_a = 591$ SLPM³) produced higher solids circulation rate. When the standpipe bed level was low (5.2 m), more aeration ($F_a = 683$ SLPM) was required to generate a given solids circulation rate (Figure 5a). This dependency of solids circulation rate on the standpipe bed height is the fundamental reason that solids circulation rate lead the standpipe aeration.

Clearly, there is a complex interaction between the solids circulation rate, riser superficial gas velocity, and the system inventory when operating a CFB unit. There are also many constraints to

² In $x \pm y$ representation, x is the mean of 300 sample measurements taken every second and y is the uncertainty of x in terms of the deviation from the standard measurement. In the case of solids circulation rate measurement, the standard measurement was chosen to be a diversion of solids for a known period of time and weighing the amount of solids collected during that time [30].

³ SLPM stands for Standard Liters per Minute

consider in order to avoid instabilities of CFBs. Decoupling solid hydrodynamics between the riser and the standpipe arising from the use of non-mechanical valve may be an attractive option that provides ease in CFB operation. A mechanical device, such as the screw feeder [46], can be used in place of the non-mechanical valve to achieve decoupling. However, it is difficult for any mechanical feeder to move large amounts of hot solids from the standpipe to the riser bottom. For example, a 165 MWe CFB boiler would need to transfer nearly 3600 tons/hr of hot solids. Continuously transferring such large amounts of hot solids will lead to mechanical deterioration because the mechanical parts have limitations in withstanding high temperatures [46].

Therefore, feedback will always present in the CFBs when the non-mechanical valve is used between the riser and the standpipe. The pressure balance (Eq. 24) and mass balance (Eq. 26) govern the feedback mechanism in the circulating fluidized beds. The pressure drops depend on the complex interaction between the riser gas velocity, the solids inventory, and the solids circulation rate [47]. Hence, the knowledge of these parameters is of paramount importance to ensure stable operation and robust control of a CFB unit.

3.4 Step Change, Modulation and Shutdown Experiments

Modeling of the large-scale CFB unit benefits from testing on steady state operation, as well as in dynamic conditions. Extensive validation builds confidence in the model use. It provides better understanding of large CFB boilers. The state model (Eq. 7) was compared against different dynamic data sets collected with HDPE, glass beads and cork particles. The transient conditions included step change, sinusoidal variation in solids flow and solids cut off experiments of HDPE beads. Only modulation experiments were considered in the case of glass beads and cork particles.

3.4.1 Solution Method

The dynamic model (Eq. 7) was solved using a 4th order Runge Kutta method. The solution method requires initial value of the riser inventory ($M_r(0)$) as well as the mean residence time (τ_r) of

solids. The average residence time of solids was determined from the slip velocity of solids with respect to the gas velocity (Eq. 14). The value of calibration parameter κ was chosen between 0.20 and 0.24 that is $\pm 2\%$ of its nominal value (0.22). Using the mean residence time (τ_r), the initial inventory was calculated from the solids circulation rate, that is, $M_r(0) = \tau_r M_s(0)$ where $M_s(0) = M_s$, a steady value of solids circulation rate prior to time 0. The time series data of solids circulation rate was obtained using the spiral velocity and the nominal voidage (ϵ_n) but not the voidage calculated from the standpipe model. In [30], it was shown that the uncertainty of mean solids circulation rate using a nominal voidage was $\pm 10\%$ of the average value of the solids flow rate measured from the standard solids diversion technique [48]. Therefore, the $\pm 2\%$ deviation in the calibration parameter κ is well within the experimental uncertainty of solids circulation rate measurement that represents $M_{s,in}$ in Eq. 7.

Generally, solids inventory in the riser is not measured. First, the estimation of solids inventory is used to calculate the pressure drop and, the results are compared against the measured values of the pressure differential. The pressure drop across the riser (ΔP_r) also depends on the frictional loss (Eq. 16). The solids and gas friction factors were calculated from Eqs. 20 and 22. In dynamic simulation, the particle velocity (u_{pfd}) required to calculate solids frictional loss was determined using $u_{pfd} = U_g/\epsilon_{fd} - u_t$, where ϵ_{fd} came from Eq. 21. Figure 4c shows that the particle velocity determined from the terminal velocity is 4% higher than the velocity (u_{pfd}) calculated from the force balance equation (Eq. 17). The single particle terminal velocity was not corrected by a factor $\epsilon_{fd}^{2.65}$ that might have resulted in the 4% discrepancy. Using the correction makes $u_{pfd} = U_g/\epsilon_{fd} - u_t \epsilon_{fd}^{2.65}$ which, in turn, makes a dynamic simulation more complex in real time applications because a numerical technique will be required to solve for u_{pfd} . In fully developed zone, the voidage is anyway closer to 1 and as a result, the drag or terminal velocity correction will be closer to 1 as discussed earlier.

3.4.2 Model Prediction

The NETL CFB unit operates the standpipe between the packed and fluidized bed regimes, namely in the moving bed condition [49]. A characteristic of a moving bed is determined by the viscous

and kinetic energy losses which depend on the low and high aeration flow rates respectively [50]. At low gas flow rates the moving solids have to overcome a large frictional force imparted by the standpipe wall. The moving bed exhibits the stick-slip motion that is more pronounced at low solids circulation rates (Figure 6a). The stick-slip motion may be inferred from the periodic spiral signal oscillating at 0.04 , 0.07 and 0.14 Hz (Figure 6b). The box whisker plot in Figure 6c indicates the skewness of solids circulation rate about its median at the step change intervals. The first interval (M_{s1}) shows a higher skewness than the solids flow rate M_{s2} and M_{s3} in the remaining two time intervals. The periodic nature of the solids circulation rate propagated to the model prediction of the riser inventory. In Figures 7a and 9a, at low M_s , these oscillations are visible in the pressure drop estimate of the riser. The solids circulation rate was changed from one level to the next to bring the changes in ΔP_r at a fixed gas velocity. In Figure 7, the dotted red line represents the riser pressure drop at $\pm 10\%$ step change in the solids circulation rate. The superficial gas velocity was kept at $U_g = 10.67$ m/s. The solid blue line represents the model prediction using the solids inventory calculated from the residence time. During transients, the model of Eq. 7 matched the pressure drop well. The step changes in the riser pressure drop from their previous steady states were adequately captured.

The objective of the modulated experiments was to investigate the phase relation between standpipe aerations and many primary response variables. The modulation was accomplished by sinusoidally perturbing the standpipe aeration by 20% up and below its steady state value. The period of the oscillations was maintained at 60 seconds. As a result, the solids circulation rate and the pressure drop signals were also modulated at the imposed 60 seconds period. Ludlow et. al. [43] determined solids circulation rate by calculating the phase angle between the standpipe aeration and the riser pressure drop. The phase angles were calculated for different transport regimes of the riser and at low (5.2 m) and high standpipe bed heights (10.7 m). Figures 8a, 8b and 8c represent the comparison of experimental and modeled pressure drop at different transport regimes of the riser. These experiments were conducted at low standpipe bed height of 5.2 m. The red dotted line indicates the experimental pressure drop and the

blue solid line represents the predicted value. The shaded area indicates experimental uncertainty of the measured pressure drop that was calculated as follows: First, the average of 5 periods were taken,

$$\begin{aligned}\Delta P_r^{ave}(1) &= \frac{\Delta P_r(1)+\Delta P_r(61)+\Delta P_r(121)+\Delta P_r(181)+\Delta P_r(241)}{5} \\ \Delta P_r^{ave}(2) &= \frac{\Delta P_r(2)+\Delta P_r(62)+\Delta P_r(122)+\Delta P_r(182)+\Delta P_r(242)}{5} \\ &\vdots \\ \Delta P_r^{ave}(60) &= \frac{\Delta P_r(60)+\Delta P_r(120)+\Delta P_r(180)+\Delta P_r(240)+\Delta P_r(300)}{5}\end{aligned}\quad (28)$$

The five period averaged values (ΔP_r^{ave}) were extended for the total duration of 5 minutes, that is,

$$\Delta P_r^{ave}(1:60) = \Delta P_r^{ave}(61:120) = \dots = \Delta P_r^{ave}(241:300) \quad (29)$$

The representation such as $1:60$ within the parenthesis indicates the time starting at $t = 1$ second and ending at $t = 60$ seconds. Then, the standard deviation (σ_r) of the measured value from its average was calculated as,

$$\sigma_r = \sqrt{\frac{(\Delta P_r(1)-\Delta P_r^{ave}(1))^2 + \dots + (\Delta P_r(300)-\Delta P_r^{ave}(300))^2}{299}} \quad (30)$$

Finally, the value of $2\sigma_r$ was added and subtracted from the experimental pressure drop as a measure of $\pm 95\%$ confidence interval. In Figure 8, we see significant fluctuations in the measured pressure drops but the prediction appears comparatively smoother due to integration of mass balance equation. On average, the model predicted the experimental ΔP_r well.

Monazam et. al. [41] introduced the solids cut off experiments to measure the solids elutriation rate and the saturation carrying capacity of the riser gas. The solids flowing into the riser, $M_{s,in}(t)$, were cut off in the standpipe after CFB came to a steady state of the underlying operating condition. A continuous flow of gas was maintained at the riser bottom. Eventually the solids left the riser letting it

completely empty after some time. As the solids inventory was carried away from the riser, the pressure drop across the riser decreased. This test was also conducted in our study and results are plotted in Figure 9. The experimental pressure drop showed a hump due to residual solids pouring into the riser from the L-valve. The control valve supplies a L-valve sparger and it takes some time to fully close the valve that might have led some solids to pour into the riser. The state model of the riser gave a correct estimation of the solids inventory leaving the riser although it could not capture the hump observed in the experiments.

3.4.3 *Effect of Initial Condition and Sensitivity of τ_r in Model Prediction*

As long as the calculated mean residence time of solids represents the average time solids spend in the riser, the effect of initial riser inventory on model prediction is minimal. Figure 10a shows the model prediction starting from $M_r(0) = 15.2$ kg, 12.2 kg, and 18.8 kg respectively. The former value is the nominal value calculated from the residence time and the latter two are $\pm 20\%$ higher than the nominal. The nominal value of τ_r was 13.8 seconds and it was varied between $\pm 10\%$ from the nominal. The data was collected using the cork bed material modulated at 60 seconds period and circulated at a high solids flow rate of 1.1 kg/s and a superficial gas velocity of 5.2 m/s. Figures 10b indicates the modulated glass beads experiments where the period of modulation was kept at 48 seconds. The steady solids circulation rate and the superficial gas velocity were maintained at 12.6 kg/s and 5.2 m/s respectively. The nominal value of initial riser inventory was 174.5 kg and it was changed by $\pm 20\%$, which is 139.6 kg and 209.4 kg. The mean residence time of solids (τ_r) was only varied by $\pm 10\%$ from the nominal value of 13.8 seconds.

In Figure 10, the dotted red line indicates the experimental data and the solid blue line is the model prediction based on nominal values of τ_r and $M_r(0)$. In both cork and glass beads data, the uncertainty of experimental measurements was calculated as in the HDPE case. The green and light blue lines represent the $\pm 20\%$ variation in $M_r(0)$ keeping τ_r at its nominal. Within the first period of modulation, the models starting from low and high riser inventory from the nominal follow the measured

pressure drop within the experimental uncertainty. Certainly, the model prediction with a lower initial value of the riser inventory starts lower than the nominal and the prediction with a higher initial value start higher. Compared to the $\pm 20\%$ variation in initial riser inventory, the $\pm 10\%$ variation in the residence time caused significant deviation in the model prediction of the experimental pressure drop. In Figure 10, the grey and black solid lines indicate low and high variation in the residence time corresponded to 12.4 and 15.2 seconds, respectively from the nominal value of 13.8 seconds. The deviation in solids residence time resulted in the pressure drop prediction that fell well outside the experimental uncertainty of the measurements. Hence, the mean residence time of solids is more sensitive compared to the initial value of the riser inventory in the model prediction. For successfully modeling the total pressure drop across the riser, the accurate knowledge of solids residence time is required that can be determined from the approach given in Section 2.

4. Conclusions and Future Work

A circulating fluidized bed operates in a closed loop. Due to a non-mechanical valve between the riser and the standpipe, feedback is inherent in the circulating fluidized bed. The hydrodynamic coupling between the standpipe and the riser was discussed in detail. The coupling arises due to the complex interaction among system inventories, solids flow rate and superficial gas velocity. The knowledge of solids inventory in the standpipe and the riser becomes critical for stable operation of the fluidized bed. The solids circulation rate and riser gas velocity determine the solids inventory and subsequently, the pressure drop across the standpipe, non-mechanical valve, and riser.

The present paper developed the dynamic model of the riser from the conservation of solids mass. The rate of change of riser inventory was determined by taking the difference between the mass flow rates of solids into and out of the riser. The dynamic model requires mean residence time to calculate the rate of solids flowing out of the riser. Initially, the residence time was determined from the average particle velocity in the fully developed region. The riser inventory calculated from this residence time under

predicted the pressure drop by 50% when compared to its measurements. In this work, the residence time was calibrated in terms of the ratio of solids concentration in the crossover to that in the riser. For HDPE beads, glass beads, and cork particles, the solids concentration in the riser outlet accounted for 22% of the average concentration of solids in the riser. The solids inventory calculated from the mass balance equation was then used to predict the overall pressure drop across the riser. The pressure drop model was derived from the momentum balance equation. Momentum balance involved contributions of the hydrostatic head of solids and the wall frictions due to gas and solids. The prediction of pressure drop was validated against the experimental pressure drop measured at different transient conditions. The dynamic experiments included step change, modulation and shutdown of HDPE flow rate as well as the modulation of flow rates of glass beads and cork particles. During experiments, the riser superficial gas velocity was kept above the upper transport velocity. In these operating conditions, the dynamic model adequately captured the hydrodynamics of the riser as a function of time.

Power plants or oil refineries use harsh operating conditions. Real time measurement of solids circulation turns out to be challenging at high temperature and elevated pressure. Even the dynamic data of solids circulation rate may not be readily available from the cold flow circulating fluidized beds. Future work entails development of a robust method for estimating solids flow rate in real time. The dynamic models of the remaining components are essential for describing the mass and pressure balance mechanisms in the transient conditions. These models, in conjunction with the present state model, may assist in simulating the full CFB loop. In addition, the entire model may help design of automatic control, troubleshooting, plant optimization, and training of operators working in CFB plants.

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Nomenclature

ΔP_{sp}	Pressure drop across the standpipe, Pa	ε_p	Packed void fraction, -
ΔP_{lv}	Pressure drop across the L-valve, Pa	ε_n	Nominal voidage in the standpipe, -
ΔP_r	Pressure drop across the riser, Pa	ε_f	Fluffed void fraction, -
ΔP_{co}	Pressure drop across the crossover, Pa	ε_r	Average voidage in the riser, -
ΔP_{cy}	Pressure drop across the cyclone, Pa	ε_{fd}	Average voidage in the fully developed zone of the riser, -
M_t	Total solids inventory, kg	ε_{co}	Average voidage in the crossover, -
M_{sp}	Solids inventory in the standpipe, kg	Δh_r	Length of the riser, m
M_{lv}	Solids inventory in the L-valve, kg	h_0	Riser base, m
M_r	Solids inventory in the riser, kg	h_{top}	Riser top, m
$M_r(0)$	Riser inventory at time 0, kg	D_r	Internal diameter of the riser, m
M_{co}	Solids inventory in the crossover, kg	A_{rs}	Cross-sectional area of the riser, m^2
M_{cy}	Solids inventory in the cyclone, kg	A_{co}	Cross-sectional area of the crossover, m^2
M_s	Solids circulation rate at steady state, kg/s	H_{sp}	Standpipe bed height during operation, m
$M_{s,in}$	Solids flowing into the riser, kg/s	Δh_{sp}	Length of the standpipe, m
$M_{s,out}$	Solids flowing out of the riser, kg/s	ΔL_{lv}	Length of the L-valve, m
F_a	Aeration in the standpipe, $SLPM$	ΔL_{co}	Length of the crossover, m
u_g	Interstitial gas velocity in the riser, m/s	ΔL_{cy}	Length of the cyclone, m
U_g	Superficial gas velocity in the riser, m/s	d_p	Particle diameter, m
U_{tr1}	Lower transport velocity in the riser, m/s	ϕ_p	Particle sphericity, -
U_{tr2}	Upper transport velocity in the riser, m/s	ρ_p	Particle density, kg/m^3
U_{mf}	Minimum fluidization velocity, m/s	ρ_g	Gas density, kg/m^3
u_t	Particle terminal velocity, m/s	μ_g	Viscosity of air, $Pa\cdot s$
u_p	Mean solids velocity in the riser, m/s	f_s	Solids-Wall friction factor, -
u_{pfd}	Solids velocity in the fully developed region, m/s	f_g	Gas-Wall friction factor, -
u_{slip}	Slip velocity of gas with respect to solids velocity, m/s	C_d	Drag coefficient, -
A_r	Archimedes Number, -	τ_r	Mean solids residence time in the riser, s
F_r	Froude Number based on U_g , -	ψ	Slip factor, -
Fr_t	Froude Number using u_t , -	κ	Normalized solids concentration in the riser outlet, -
Re_g	Reynolds Number using U_g , -	σ_r	Standard deviation of the modulated riser pressure drop, Pa
Re_t	Reynolds Number using u_t , -	t	Time, s
Re_s	Reynolds Number using $U_g/\varepsilon_{fd}-u_{pfd}$, -	f	Frequency, Hz
		g	Acceleration due to gravity, m/s^2

Disclaimer

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Figures

Figure 1. Schematic of the NETL circulating fluidized bed unit. Green arrows represent gas flow and the light blue arrows indicate the solids flow. The aeration F_{a0} is referred to as an educator that is occasionally used in high solids circulation rate experiments. The secondary cyclone separates the residual solids leaving the primary cyclone above the standpipe. The separated solids are fed back to the standpipe by opening the valve (not shown) or are sent directly to the bag house.

Figure 2. Comparison of the pressure drop prediction against the experimental pressure drop. The prediction was calculated based on the slip ratio correlation proposed by Patience et. al. [16].

Figure 3. Apparent axial voidage profile in the riser operating at the homogeneous transport regime. Symbols represent actual experimental data and lines represent the polynomial fit of the apparent voidage ($\epsilon_{apparent}$ (-)) as a function of incremental height (h_i (m)).

(a) Independent values of parameter κ for three different bed materials including HDPE particles, glass beads, and cork particles in the riser.

(b) Determination of a single parameter κ for HDPE, glass beads, and cork particles in the riser.

- (c) Comparison of the average particle velocity in the fully developed zone

Figure 4. Calibration of a parameter κ from the mass inventory of HDPE particles, glass beads and cork particles in the riser. In addition, the average particle velocity (u_{pfd}) derived from Eq. 17 in the fully developed zone and that derived from the terminal velocity (u_t) are shown.

(a) M_s leads F_a

(b) Low H_{sp}

(c) M_s in phase with F_a

(d) High H_{sp}

Figure 5. Sinusoidal F_a , M_s and H_{sp} at low and high system inventories of HDPE beads. The bed height, H_{sp} , is 180° out of phase from the standpipe aeration, F_a .

(a) Time domain solids circulation rate

(b) Frequency domain solids circulation rate

(c) Box whisker plot of solids circulation rate at step change intervals

Figure 6. Illustration of a stick-slip motion of HDPE solids in the standpipe using the spiral signal.

(a) $U_g = 10.70$ m/s and $M_s = 1.25$ kg/s

(b) $U_g = 10.73$ m/s and $M_s = 6.44$ kg/s

(c) $U_g = 10.68$ m/s and $M_s = 12.26$ kg/s

Figure 7. Step change experiments with HDPE particles where the solids circulation was changed $\pm 10\%$ from the steady state condition.

(a) $U_g = 10.75$ m/s and $M_s = 1.25$ kg/s

(b) $U_g = 10.43$ m/s and $M_s = 6.83$ kg/s

(c) $U_g = 10.45$ m/s and $M_s = 12.82$ kg/s

Figure 8. Modulated experiments with HDPE beads where the solids circulation was modulated at 60 seconds period with 20% perturbation from the steady state value.

(a) $U_g = 10.70$ m/s and $M_s = 1.25$ kg/s

(b) $U_g = 10.73$ m/s and $M_s = 6.44$ kg/s

(c) $U_g = 10.68$ m/s and $M_s = 12.26$ kg/s

Figure 9. Shutdown experiments with HDPE beads where the solids circulation was cutoff but the riser gas flow was maintained to empty out the riser.

(a) Modulated experiments with cork particles where the solids circulation was modulated at 60 seconds period

(b) Modulated experiments with glass beads where the solids circulation was modulated at 48 seconds period

Figure 10. Effect of initial riser inventory, $M_r(0)$, and the solids residence time, τ_r , on model prediction of the riser pressure drop.

Table 1. Granular material properties of three different solid particles used in this study.

Bed Material		<i>HDPE</i>	<i>GB</i>	<i>Cork</i>
Geldart Group		Group B	Group A	Group B
Particle Diameter, d_p	μm	802	59	812
Particle Sphericity, ϕ_p	-	0.95	0.98	0.84
Particle Density, ρ_p	kg/m^3	863	2425	189
Void Fraction Fluffed, ε_f	-	0.410	0.421	0.440
Void Fraction Nominal, ε_n	-	0.350	0.410	0.430
Void Fraction Packed, ε_p	-	0.346	0.371	0.370
Bulk Density Fluffed, $\rho_p(1-\varepsilon_f)$	kg/m^3	509	1404	106
Bulk Density Packed, $\rho_p(1-\varepsilon_p)$	kg/m^3	564	1526	119
Minimum Fluidization Velocity, U_{mf}	m/s	0.13	0.03	0.09
Lower Transport Velocity, U_{tr1}	m/s	4.33	1.84	2.30
Upper Transport Velocity, U_{tr2}	m/s	6.25	2.96	3.59
Particle Terminal Velocity, u_t	m/s	2.89	0.22	0.95
Archimedes Number, $Ar = g d_p^3 \rho_g (\rho_p - \rho_g) / \mu_g^2$	-	15811.6	17.6	3574.8
Reynolds Number, $Re_t = \rho_g u_t d_p / \mu_g$	-	153.1	0.9	51.1
Froude Number, $Fr_t = u_t / \sqrt{g D_r}$	-	1.6	0.1	0.5
d_p/D_r	-	2.6×10^{-3}	1.9×10^{-4}	2.7×10^{-3}

‡ Gas density (ρ_g) and air viscosity (μ_g) are in Std. conditions.

Graphical Abstracts