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Studies on solid mean residence time in a three-stage gas-solid fluidized bed with downcomer

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Abstract−Fluidized bed reactors behave as a continuously stirred tank reactor having wide residence time of solids, which is not desirable if a homogeneous product is required. The multi-stage fluidized bed reactors narrow the solids residence time, making it useful for various operations. A three-stage fluidized reactor was designed, fabricated and operated under stable operating condition to investigate the mean particle residence time in the system. The materials taken for the study were lime and sand. In the particle residence time experiments, the results revealed that at a particular solids velocity, mean residence time decreased with increase in gas velocity and increased with decrease in gas velocity. Based on the data, a correlation has been presented for predicting mean residence time.

Key words: Multi-stage, Fluidization, Perforated Plates, Mean Residence Time, Downcomer

INTRODUCTION

Fluidization has found very successful applications in gas-solids two-phase operations [1-4]. Gas-solids fluidization is one of the most important techniques adopted in industry due to the favorable gas-solids contacting areas. However, considering the flow of solids, a continuous fluidized bed reactor behaves as a continuously stirred tank reactor and the residence time of solids is very wide, which is not desirable if a homogeneous product is required. An alternate way to narrow the solids residence time is to use a multistage fluidized bed reactor (MFBR). In a common MFBR, particles are fed from the top of the column and depart from the bottom of the column; therefore, gas-solids flow counter-currently. Perforated grid plates are used to section the column into several stages. The addition of perforated plates has been reported to narrow residence time distribution [2,5], eliminate slugs, break bubbles and agglomerates, thus realizing stable operation, with minimal solids back-mixing, and uniform axial voidage distribution. The staged fluidized bed can be operated co-currently or counter-currently, with or without downcomers, and with or without the recycle of solids [6]. Mostly, the counter-current multistage fluidized bed reactors are preferred due to their easy operation. The literature also suggests that co-current multi-stage fluidized bed reactor has been applied for the biomass gasification and showed improved gasification efficiency compared to conventional non-staged operations [7]. A number of correlations have been proposed to predict the pressure drop across a multi-stage "bubbling" type bed [9,10]. The correlations are complex functions of Reynolds number, Archimedes number, the gas velocity, the plate design and the physical properties of the particles. The influences of fluidized bed hydrodynamic parameters at high temperature have been reported to understand the solid attrition [11].

Common to all counter-current multi-stage fluidized bed systems, there are two major operating difficulties: the maintenance of a stable bubbling operation and the prediction of the particle residence time for a given system. Determining the values of the operating parameter for a stable bubbling operation and the value of the particle residence time requires an accurate understanding of the influence of each operating parameters on the flow regimes as well as an accurate model of system's hydrodynamics. Apparently, the stable operating ranges are different for different particles in different studies. This is because the particles and the system designs can be very different for different applications, and it is almost impossible to have a complete understanding of the influences of every operating parameter on the bed behavior. A counter-current multi-stage fluidized bed system has been reported useful in drying, adsorption, heat recovery and other applications. Because each stage being identical in its performance to the rest of the stages in multi-stage fluidized bed systems [1-3,8], we focused on the particle residence time of a counter-current three-stage fluidized bed system with downcomers in the present work. The particle residence time in a counter-current staged bed is affected by the discharge of solids through the downcomer, gas flow rate, opening ratio of the plate, pressure fluctuations in the fluidized bed and the plate designs. Since the system designs are usually different for different applications, it is almost impossible to relate the particle residence time to every operating parameter. Therefore, we focused on most important parameter i.e. effect of solids flow rate on particle residence time of a three-stage fluidized bed system with downcomer.

EXPERIMENTAL SETUP AND TECHNIQUE

Fig. 1 shows the apparatus used in the experiments. The cylindrical column made of perspex was 305 cm in length and 100 mm

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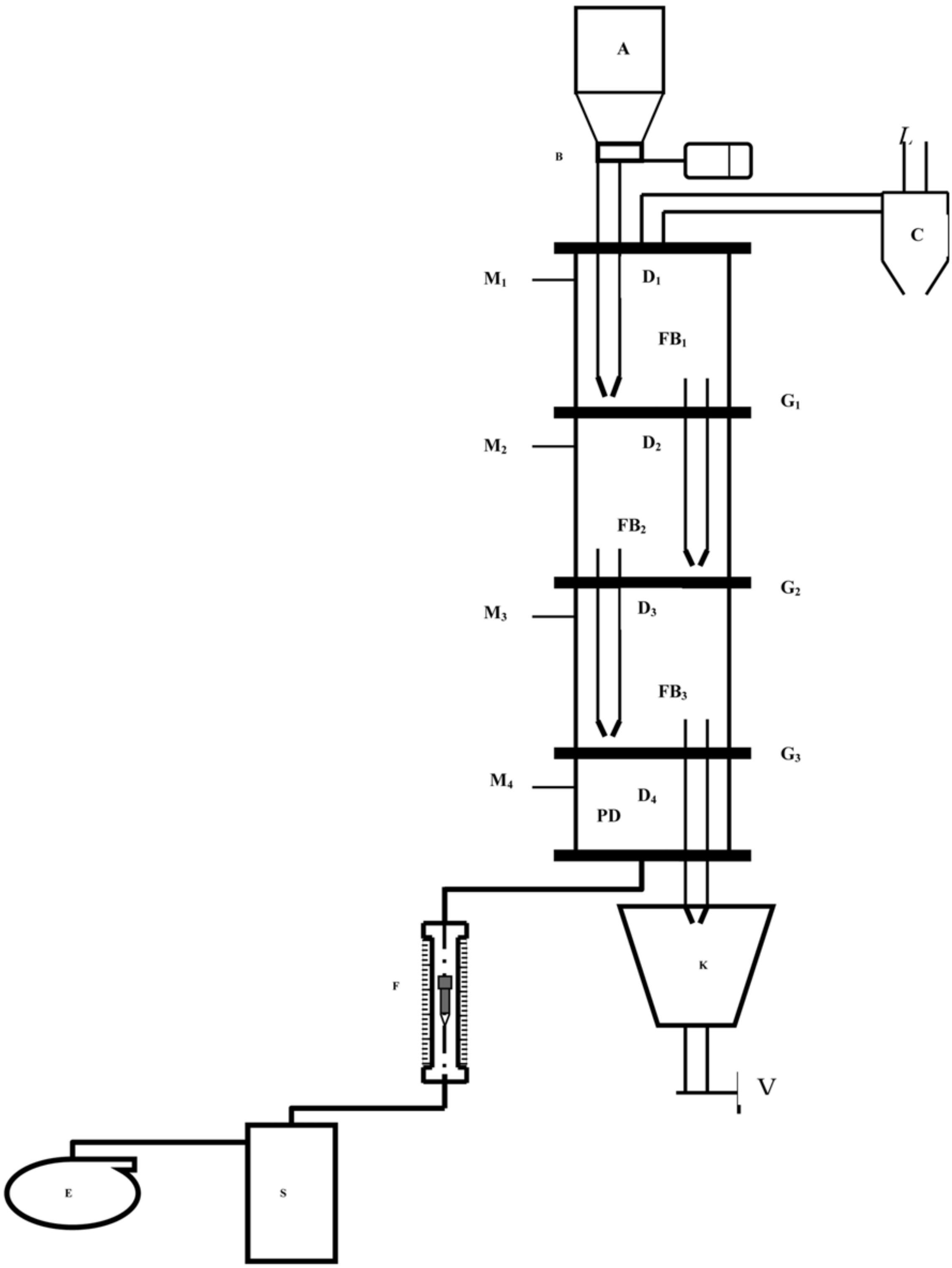

Fig. 1. Schematic diagram of the experimental set-up of a three-stage counter- current fluidized bed reactor.

A. Solids feed hopper	E. Compressor	K. Solid outlet storage	V_{1-2}. Valve
B. Screw feeder	S. Surge tank	L. Air outlet	
C. Cyclone	F. Rotameter	PD. Pre-distributor	
D_{1-4}. Downcomers	G_{1-3}. SS perforated plate	M_{1-4}. Manometers	

in internal diameter. Air was introduced to the system positively from a tank connected to a compressor. The gas flow rate was controlled by a calibrated rotameter and the superficial gas velocity was controlled based on the hydrodynamic study. Solids were fed from the top of the column into the system using a screw feeder. A distributor was used to uniformly distribute the particles on the entire cross-section area of the column. Stainless steel plates of 0.002 m thick each (G_1, G_2 and G_3) were used as internal baffles between two stages and each plate was drilled with perforations of 0.002 m diameter on a triangular pitch having 8.56% total grid openings. The grid plates were covered with fine wire mesh (100 mesh size) to prevent the solids from falling through the openings. Each section was provided with a downcomer of perspex cylinder of 0.025 m internal diameter (D_1, D_2, D_3 and D_4), and the downcomers were fitted to the gas distributors by special threading arrangement having the provision for adjusting the weir height as desired. The downcomers were further fitted with a cone at the exit end in order to reduce the up flow of the gas through the downcomer and, consequently, widening the stable operating range. The bottom-opening diameter of the cone was 0.012 m for flow of solids to the next stage. Two identical plates were used in the experiments. Pressure tapings were provided just below the grid plate and the near the air out let and four manometers were provided to measure the pressure drop at every stage as well as the total pressure drop. The pressure drops across the upper and the lower stages were measured independently. The gas leaving the column from the top stage was passed through a 0.15 m diameter standard cyclone (C) and then into the exhaust system. A cloth bag was attached at the bottom of the cyclone to collect the fines,

Table 1. Properties of bed materials

Material	Density (ρ_p) kg/m^3	Particle size (d_p), μm	Static bed porosity (ε_{mf})	Sphericity (ψ)	U_{mf}, m/s
Lime	2040	426	0.48	0.7	0.112
Sand	2650	426	0.48	0.76	0.135

if any, carried over.

The hydrodynamic data has been taken into consideration for stable operation of reactor [1-3]. Table 1 lists the physical properties of materials studied, i.e., lime particles and sand particles. The solids were fed into the column at the rate from 0.035 kg/m^2·s (1.0 kg/hr) to 0.142 kg/m^2·s (4.0 kg/hr) with gas flow rate varying from 0.32 (0.265 m/s) to 0.59 kg/m^2·s (0.49 m/s). The weir height of the downcomer was kept constant at 50 mm and the gap between the downcomer bottom and the grid plate were kept 25 mm. When the system was operated with solids, it was observed that all the stages of the reactor were identical in their operation as well as performance. No discernible difference in the solids hold-up across the stage was noticed from stage to stage. Once the system was allowed to reach equilibrium or steady state, the gas and the solids flow was then cut off simultaneously and the solids were weighed to calculate the mean residence time. Since the solids holdup and mean residence time are mutually interrelated, the mean solids holding time, t, in one stage is defined as follows:

$$t = \frac{\text{Solids holdup in the stage (W)}}{\text{Solids flow rate, m (=AGs)}} \tag{1}$$

RESULTS AND DISCUSSION

1. Effect of Solids Velocity on Solids Hold-up

The steady state holdup for all the stages was studied as a function of the solids flow rate. Fig. 2 describes the effect of solids flow rate on solids holdup measured across each stage varying the solid flow rates from 16.67 g/min to 66.8 g/min in the three-stage reactor operated under steady-state conditions, with the solids particles and gas flowing counter-currently. It is seen from the figures that

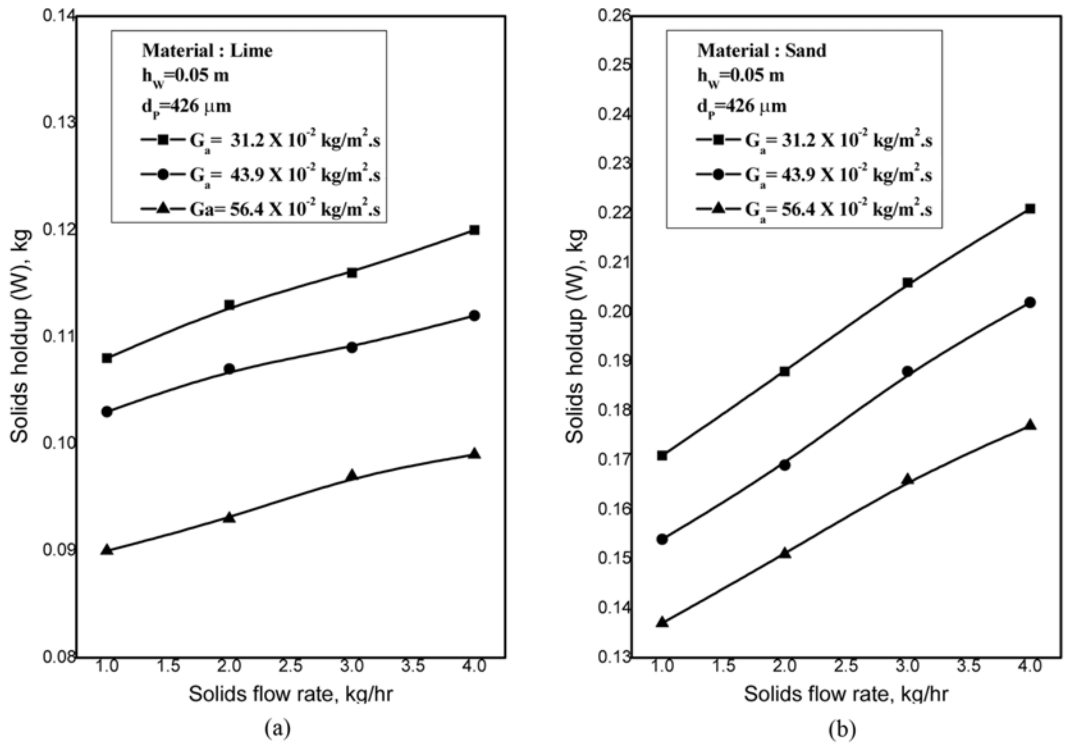

Fig. 2. Effect of solids flow velocities on solids holdup for (a) lime (b) sand particles.

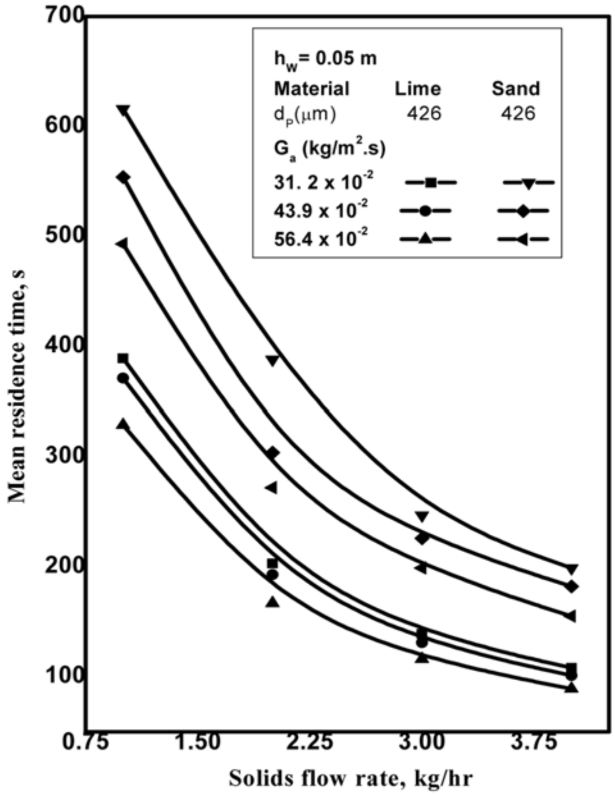

Fig. 3. Effect of solids flow rate on mean residence time for lime and sand particles.

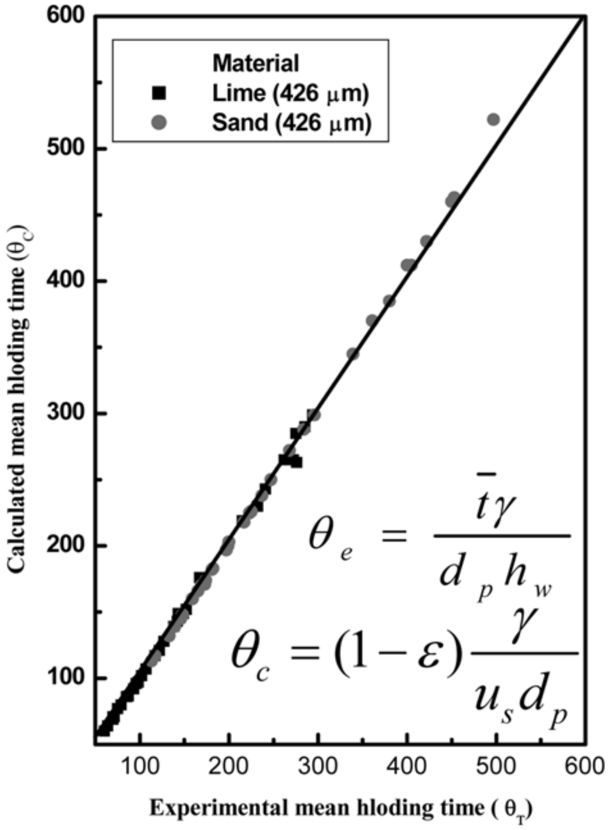

Fig. 4. Comparison of experimental values of mean solids holding time (θ_e) with that calculated mean solids holding time (θ_c).

the solids holdup, W, decreases with increase in the gas flow rate and increases with increase in the solids flow rate. The minimum solids holdup obtained in the column at high gas flow rate (0.58 kg/m^2·s) corresponding to minimum solid flow rate (0.035 kg/m^2·s) is 0.09 and 0.137 kg at 0.05 m weir height for lime and sand particles, respectively, and the maximum solids holdup obtained in the column at low gas flow rate (0.32 kg/m^2·s) corresponding to maximum solid flow rate (0.141 kg/m^2·s) is 0.12 and 0.221 kg at 0.05 m weir height for lime and sand particles, respectively. The reason may be that an increase in gas rate increases the porosity of the bed in the system, resulting in a decrease in the solids concentration and hence the solids holdup at the stage.

2. Effect of Solids Velocity on Mean Residence Time

Fig. 3 describes the effect of solids flow rate on mean particle residence time. When the system was operated with solids, it was observed that all the stages of the reactor were identical in their operation as well as performance. No discernible difference in the solids hold-up across the stage was noticed from stage to stage. Once the system was allowed to reach equilibrium or steady state, the gas and the solids flow was then cut off simultaneously and the solids of each stage were weighed and found to be almost the same. The solids of one stage were taken to calculate the mean residence time. Since the solids holdup and mean residence time are mutually interrelated, the mean solids holding time, t, in one stage was defined as the ratio of solids holdup in the stage to the solids flow rate. At steady state condition, the particle mean residence time for the three stages remains the same when other operating parameters are kept the same. At a particular gas velocity, the mean residence time decreased with increase in solids flow rate as Eq. (1) suggests. The mean time also decreased with the increase of the gas velocity. Since increasing the gas velocity increases the bed porosity, it decreases the residence time of particles in the bed. The solids mean residence time varied from minimum 89 sec to maximum 389 sec in case of lime particles and minimum 155 sec to maximum 616 sec in case of sand particles at the operating conditions. The Eq. (1) is satisfactorily correlated by defining a modified dimensionless holding time, θ [12].

$$\theta_e = \frac{\bar{t}\gamma}{d_p h_w} \tag{2}$$

$$\theta_c = (1-\varepsilon)\frac{\gamma}{u_s d_p} \tag{3}$$

Fig. 4 shows the correlation for mean holding time in multistage fluidized beds. The fractional solids concentrations $(1-\varepsilon)$ in the Eq. (3) depend on corresponding gas and solids flow rate and the kinematic viscosity of gas has been taken at 25 ºC. Eq. (3) was developed to calculate the mean particle residence time theoretically, considering fractional solids concentration, weir height and solids velocity. The fractional solids concentration has been considered based on a gas flow rate corresponding to each solids flow rate.

For each experiment (i.e., at a particular gas and solids flow rate), the system was allowed to reach equilibrium so that the inlet and outlet solids flow rate were equal. All the stages of the reactor were identical in their operation as well as performance. The gas and the solids

Table 2. Values of solids concentrations for lime and sand particles at 50 mm weir height

Gas flow rate (kg/m^2·s) / Solids flow rate (kg/m^2·s)	Solids concentration $(1-\varepsilon)$ Lime			Solids concentration $(1-\varepsilon)$ Sand		
	31.2×10^{-2}	43.9×10^{-2}	56.4×10^{-2}	31.2×10^{-2}	43.9×10^{-2}	56.4×10^{-2}
53.4×10^{-3}	0.134	0.127	0.112	0.163	0.149	0.133
71.0×10^{-3}	0.138	0.132	0.115	0.180	0.163	0.144
106.2×10^{-3}	0.141	0.134	0.119	0.199	0.178	0.159
141.5×10^{-3}	0.144	0.138	0.122	0.212	0.190	0.173

flow was then cutoff simultaneously and the solids were weighed [1]. The quantitative information on bed porosity/solids concentration has been calculated using the following Eq. (4).

$$1-\varepsilon=\frac{W/A}{h_w\rho_s} \tag{4}$$

Where W=amount of solids retained at a particular solids flow rate and gas flow rate
A=cross sectional area of reactor
h_w=height of weir
ρ_s=density of solids

The values of solids concentrations at different gas velocity and solids velocity for lime and sand particles at 50 mm weir height are given in Table 2.

The kinematic viscosity of gas was taken at 25 ºC and its numerical value is nearly 1.5×10^{-3} m^2/s. The physical meaning of dimensionless mean residence time is the ratio of time to equilibrium time and it's a measure of holding time irrespective of vessel size and other variables.

CONCLUSIONS

The solids holdups vis-a-vis particle residence time in the countercurrent three-stage fluidized bed reactor at a particular weir height were studied. The system was operated in a continuous fashion and had moderate particle residence time (few minutes). Under stable operating condition, standard bubbling regime was observed in the bed. The minimum solids holdup obtained in the column at high gas flow rate (0.58 kg/m^2·s) corresponding to minimum solid flow rate (0.035 kg/m^2·s) was 0.09 and 0.137 kg for lime and sand particles, respectively, and the maximum solids holdup obtained in the column at low gas flow rate (0.32 kg/m^2·s) corresponding to maximum solid flow rate (0.141 kg/m^2·s) was 0.12 and 0.221 kg for lime and sand particles, respectively. The solids mean residence time varied from minimum 89 sec to maximum 389 sec in case of lime particles and minimum 155 sec to maximum 616 sec in case of sand particles at the operating conditions. The solids holdup and mean solids residence time decreased with increase in the gas flow rate.

NOMENCLATURE

A : cross-sectional area of the column [m^2]
d_p : particle diameter [μm]
G_a : mass velocity of gas [kg/m^2·s]
G_S : mass velocity of solids [kg/m^2·s]
g : gravitational acceleration [m/s^2]
h_W : weir height [m]
t : mean particle residence time [s]
u_s : superficial solids velocity [m/s]
W : solids hold-up in the bed at steady state [kg]

Greek Letters

$(1-\varepsilon)$: fractional solids concentration
θ_c : calculated dimensionless solids mean residence time
θ_e : experimental dimensionless solids mean residence time
γ : Kinematic viscosity [m^2/s]

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