Study of Residence Time Distribution of Rough Rice in a Plug Flow Fluid Bed Dryer

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Abstract

Due to the motion of gas-bubbles and the particles back-mixing, the particles flow pattern through the plug flow fluid bed dryer deviates from the ideal plug flow and is considered as a dispersed plug flow. In this study, the residence time distribution (RTD) characteristics, flow pattern, and dispersion coefficient of rough rice in a plug flow fluid bed dryer under various experimental conditions were investigated. The effects of solid feed rate (46, 96, and 135 g/min), weir height (5 and 10 cm), and inlet solid moisture content (10 and 30% d.b) on RTD were studied using the pulse input of tracer (dye-coated rough rice). The flow of particles in the dryer is a dispersed plug flow with large deviation from ideal plug flow. The values of dispersion coefficient range from 2.60×10^{-4} to 4.50×10^{-4} m^2/s over the investigated conditions. The flow in the dryer approaches plug flow with increase in the solid feed rate, decrease in the weir height, and increase in the inlet solid moisture content.

Keywords: *Dispersion, Plug Flow, Residence Time, Rough Rice, Fluid Bed Dryer*

1. Introduction

Fluid bed dryers are used extensively for drying of wet particulate and granular materials such as agricultural products, foods, chemicals, ceramics, minerals, and pharmaceuticals. Fluid bed systems give important advantages such as good material mixing, high rates of heat and mass transfer, low capital and maintenance costs, and ease of control and material transportation [1].

In ideal situation, the flow patterns in continuous fluid bed dryers based on RTD can be considered as ideal well-mixed flow and ideal plug flow. Significant for a well-mixed fluid bed dryer is that the spread in the particles residence time within the dryer is very large, ranging from zero to infinity. Thus in an ideal well-mixed fluid bed dryer, the material within the bed is actually a mixture of particles with different drying age as well as outlet material. The ideal plug flow is characterized by the fact that the flow of material through the bed is orderly with no element of material overtaking or mixing with any other element ahead or behind. Actually, there may be lateral mixing of particles in this flow; however, there must be no mixing or diffusion along the flow path. Thus in an ideal plug flow fluid bed dryer, the outlet particles have the same residence time which is favorable for drying of products where an even product quality is important [1, 2].

In practice, the flow pattern can be quite different from the ideal plug flow. Due to the mixing effect of the gas bubbles in fluidized bed systems, the particles are mixed in axial direction and the result is passage of particles in different routes. It is evident that particles taking different routes through the bed, take different time duration to pass through the dryer. Therefore the RTD or the distribution of the exit time for particles leaving the dryer will be characteristic for dispersion in the dryer. This phenomenon is referred to as axial dispersion. The differential equation representing this phenomenon is as follows [2]:

$$\frac{\partial C}{\partial \theta} = \left(\frac{D}{uL}\right)\frac{\partial^2 C}{\partial^2 z} - \frac{\partial C}{\partial z} \qquad (1)$$

where C is the concentration (mol/m^3), D is the dispersion coefficient (m^2/s), u is the axial solid flow velocity (m/s), L is the bed length (m), and parameters z and τ are, respectively, the length coordinate (m) and the normalized time as follows:

$$z = ut + x/L \qquad (2)$$

$$\tau = tu/L \qquad (3)$$

where x is the bed length coordinate (m) and t is the time (s).

The dimensionless group (D/uL), known as the vessel dispersion number, is the parameter that measures the extent of axial dispersion. When the dispersion number is zero, the flow is ideal plug flow; for values of dispersion number less than or equal to 0.01, the solid flow through the bed can be considered approximately as plug flow with small dispersion, and values of dispersion number more than 0.01 show a large deviation from ideal plug flow, i.e., the dispersed plug flow. The values of D and D/uL can be calculated based on the parameters of the RTD such as mean residence time and variance of distribution [2].

An evaluation of the literature shows that there are few studies concerning the dispersion coefficient in the plug flow fluidized bed dryers. Reay [3] conducted an experimental study of the RTD in a plug flow fluidized bed dryer and proposed an equation for calculation of the dispersion coefficient based on the experimental data, as follows:

$$D = 3.71 \times 10^{-4} \, \frac{(U_0 - U_{mf})}{U_{mf}^{1/3}} \qquad (4)$$

where U_0 and U_{mf} are the superficial fluidization velocity and the minimum fluidization velocity, respectively, both in (m/s). As shown in this equation, the dispersion coefficient is dependent on the difference between U_0 and U_{mf}, which actually is a measure of the bubble size in the fluidized bed systems [4, 5].

Nilsson [6] investigated the drying process of the granular materials in a vibrated plug flow fluid bed dryer both experimentally and theoretically, and also investigated the effect of operating parameters such as solid feed rate, bed height, and superficial fluidization velocity on the RTD to obtain the dispersion coefficient in the dryer. The proposed equation to calculate the dispersion coefficient by Nilsson [6] is as follows:

$$D = \frac{1.49\left[0.01(H_{bed} - 0.05) + 0.00165\,\rho_g(U_0 - U_{mf})\right]u^{0.23}}{U_{mf}^{1/3}} \qquad (5)$$

where H_{bed} is the bed height (m) and ρ_g is the gas density (kg/m^3). The minimum bed height used in the Nilsson's experiments is 0.05 m, thus there is a (H_{bed}−0.05) term in the above equation.

Nilsson and Wimmerstedt [7] investigated the RTD and the dispersion coefficient of some granular materials in a pilot scale vibrated plug flow fluid bed dryer and also used the Nilsson's equation to calculate the dispersion coefficient. For one of the investigated materials, apatite (a mineral used by biological micro-environmental systems), the values of dispersion coefficient determined through the RTD measurement experiments were different from those calculated by the Nilsson's equation. It was mentioned that one explanation to this difference might be that the employed solid feeding procedure was different from that considered in developing the mentioned equation.

Satija and Zucker [8] investigated the effects of operating parameters such as vibration amplitude, fluidization velocity, and solid feed rate on the RTD of a material with relatively narrow size distribution with mean diameter of 720 μm in a pilot scale vibrated plug flow fluid bed dryer, and also presented the values of dispersion coefficient under different conditions. Also the effect of baffles provided inside the dryer on the RTD characteristics was studied. The vibration amplitude had the most significant effect on the mean residence time and the dispersion coefficient. Increasing the vibration amplitude decreased the mean residence time and increased the dispersion coefficient. The effect of fluidization velocity on the mean residence time was insignificant and the dispersion coefficient was less sensitive to the fluidization velocity as compared to the vibration amplitude. The mean residence time and the dispersion coefficient were decreased by increasing solid feed rate.

Han, et. al., [9] conducted an experimental study to determine the effects of vibration intensity, fluidization velocity, and solid feed rate on the RTD of wheat and pharmaceutical BYN granules in a pilot scale vibrated plug flow fluid bed dryer. It was found that the vibration intensity has a dominant effect on the mean residence time and the dispersion coefficient. Increasing the vibration intensity decreased the mean residence time and increased the dispersion coefficient. The effect of fluidization velocity on the mean residence time and the dispersion coefficient was not as significant as that of vibration intensity. With increase in solid feed rate, the mean residence time and the dispersion coefficient decreased and increased, respectively.

Fyhr, et. al., [10] and Wanjari, et. al., [11] conducted the mathematical modeling of the plug flow fluid bed drying process by considering the dispersion coefficient obtained using the Nilsson's equation and the Reay's equation, respectively.

It should be mentioned that the dispersion coefficient is not a very accurate measurable parameter like mass or temperature. It is highly dependent on dryer configuration (bed geometry), material properties, and operating conditions. Thus neither of the proposed equations can guarantee very good repeatability and accuracy when applied to another conditions, and it is necessary to measure the dispersion coefficient under the desired conditions. Thus the objectives of this study are, to: (a) investigate the particles flow pattern inside the dryer, (b) study the effects of various operating parameters such as solid feed rate, weir height, and inlet solid moisture content on RTD characteristics, and (c) determine the extent of dispersion coefficient of the rough rice in the plug flow fluid bed dryer.

2. Materials and Methods

2.1. Materials

For all the RTD measurement experiments long grain rough rice (Mazandaran, Iran) was used. The particles were cleaned manually and foreign matters such as stones and straw were removed. A highly water soluble black dye was used to coat the grain sample, and drying of this coated sample was done in the sun and used as the tracer for RTD measurement experiments. Table 1 lists the values of some physical and hydrodynamics properties of the rough rice determined based on the methods proposed by Mohsenin [12] and Yang [5], respectively.

Table 1. Some Physical Properties of the Rough Rice used in this Study

Property	Value
Dry particle density (kg/m3)	857
Bulk density corresponding to 10 and 30% d.b. (kg/m^3)	485 and 483
Particle equivalent diameter (m)	3.43×10^{-3}
Porosity	0.48
Minimum fluidization velocity (m/s)	1.7
Terminal velocity (m/s)	6.7

2.2 Plug Flow Fluidized Bed Dryer

A schematic diagram of the laboratory scale plug flow fluid bed dryer used for RTD measurement experiments is shown in Figure 1.

The fluidizing medium (gas) at room conditions was supplied by a centrifugal blower and heated by a controllable electrical heater before entering the plenum. The baffle plates were installed vertically inside the plenum to change the direction of the gas flow 90 degrees into the upward direction as well as induce a uniform gas velocity across the entire bed cross section. The large enough distance from top of the baffle plates to the distributor about 0.4 m prevents the gas from preferentially passing through the middle of the distributor. The distributor was a 2 mm thick perforated steel plate with 2 mm diameter holes on a 5 mm triangular pitch (open area = 0.15). The width and length of the drying vessel was 8 and 100 cm, respectively. For observation of hydrodynamics of the gas-solid fluidization, a tempered glass viewing section was incorporated on the side of the drying vessel.

The particles were fed from the bin to the dryer by a screw conveyor. Due to the fluidization, particles move axially along the dryer and exit from the outlet. For the output, a manually controllable sliding weir was used. For adding the tracer into the bed, a tracer inlet port was incorporated on the side of the drying vessel and near the solid inlet port. The inlet gas flow rate was adjusted by changing the speed of the centrifugal blower with an inverter (LS, SV040iG5-4, Korea). The superficial fluidization velocity was calibrated via a pitot tube (Kimo, TPL-03-200, France) and a differential pressure transmitter (Sensys DPUH 1000, Korea). The solid feed rate was adjusted by changing the rotational speed of the screw conveyor with a three phase electric motor equipped with an inverter (LS, SV004iG5-4, Korea).

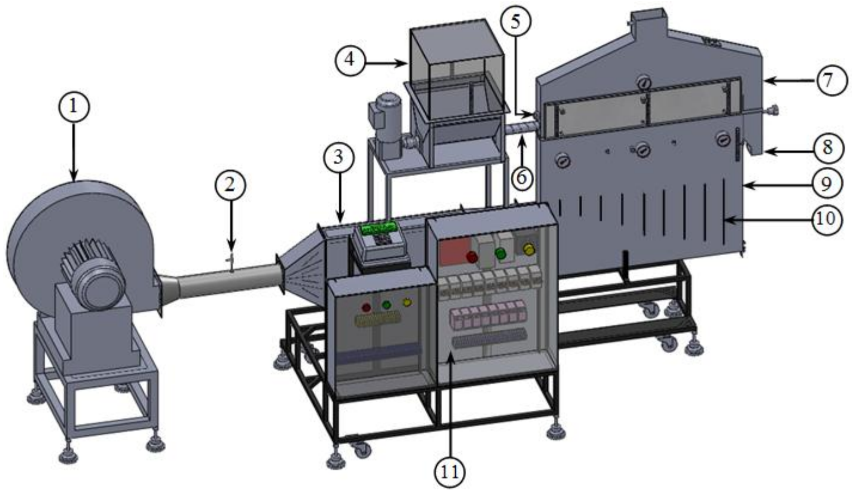

Figure 1. Schematic Diagram of the Laboratory Scale Plug Flow Fluidized Bed Dryer: (1) centrifugal blower; (2) pitot tube; (3) electrical heater; (4) bin; (5) tracer inlet port; (6) screw conveyor; (7) drying vessel; (8) solid outlet port; (9) plenum; (10) baffle plate; (11) control unit

2.3 Determination of RTD

In this study, the RTD curves were determined using the pulse stimulus-response technique. After the dryer was brought to steady state condition, i.e., the constant outlet solid mass flow rate, the quantity of the tracer approximately equal to 8 weight percent of the solid holdup within the bed was added through the tracer inlet port at zero time and a timer was started. The output samples were collected every one minute after adding the tracer until the tracer disappeared in the collected samples. Every experiment was repeated two times at the specific operating condition, and the mean values of investigated parameters were considered.

2.4 Residence Time Distribution Analysis

The exit concentration curve of the tracer C(t), the exit age distribution or RTD function E(t), and the accumulated RTD function F(t) are often used to describe the RTD characteristics. In this study, C(t) curve was obtained by plotting the tracer concentration, i.e., the mass fraction of the tracer in the output samples against the time. By normalizing C(t) curve, E(t) function is as follows [2]:

$$E(t) = \frac{C(t)}{\sum_0^\infty C(t)\Delta t} \qquad (6)$$

The parameters of RTD or E(t), such as the mean residence time (τ_R: average time of passage, or when the curve passes by the exit) and the variance of distribution (σ^2: a measure of the spread of the curve) are calculated using Equations (7) and (8), respectively [2].

$$\tau_R = \frac{\sum_0^\infty t\, C(t)\Delta t}{\sum_0^\infty C(t)\Delta t} \qquad (7)$$

$$\sigma^2 = \frac{\sum_0^\infty t^2 C(t)\Delta t}{\sum_0^\infty C(t)\Delta t} - \tau_R{}^2 \qquad (8)$$

The cumulative residence time distribution function F(t) is [2]:

$$F(t) = \sum_0^t E(t)\Delta t \qquad (9)$$

The F(t) function based on the dimensionless time variable (θ) can also be expressed as the cumulative residence time distribution F(θ) as follows [2]:

$$F(\theta) = F(t) \qquad (10)$$

where the dimensionless time variable (θ) is the ratio of time (t) to the mean residence time (τ_R) as follows [2]:

$$\theta = t/\tau_R \qquad (11)$$

The linear residence time (τ_H) can also be obtained using solid holdup within the bed (W) and solid feed rate (F) as follows [2]:

$$\tau_H = W/F \qquad (12)$$

At steady state, the flow of particles was stopped suddenly and simultaneously the collection of the particles at the dryer outlet was started until the emptying of entire material from the bed. The weight of collected sample was the solid holdup within the bed.

The flow behavior at the inlet and outlet ports of the dryer strongly affects the RTD as well as the relationship between RTD parameters and D/uL. In this study, a closed-closed boundary condition is employed, i.e. the flow of particles outside the dryer up to the inlet and outlet boundaries is plug and inside the dryer is dispersed [2].

The parameter D/uL can be related to mean residence time and variance of distribution by trial and error using the following equation [2]:

$$\sigma_D{}^2 = \frac{\sigma^2}{\tau_R{}^2} = 2\left(\frac{D}{uL}\right) - 2\left(\frac{D}{uL}\right)^2 \left[1 - e^{-uL/D}\right] \qquad (13)$$

3. Results and Discussion

3.1 RRD Characteristics

Table 2 shows the details of the experiments carried out in the plug flow fluid bed dryer to study the effects of solid feed rate, weir height, and inlet solid moisture content on the RTD. The RTD measurement experiments were conducted at solid feed rates of 46, 96, and 135 g/min, weir heights of 5 and 10 cm, and inlet solid moisture contents of 10 and 30% d.b. All the experiments were conducted at superficial fluidization velocity of 2.5 m/s under ambient temperature.

Table 2. Operating Conditions and Results of the RTD Analyses

Run	F (g/min)	H (cm)	Min (%, d.b.)	W (kg)	τ_R (min)	σ^2 (min^2)	$\sigma_D{}^2$ (-)	τ_H (min)	D/uL (-)	D×10^4 (m^2/s)
1	46	5	10	0.685	12.48	55.28	0.3549	14.89	0.232304	2.60
2	46	10	10.1	1.330	21.55	302.57	0.6515	28.91	0.683504	3.94
3	96	5	9.9	0.740	6.25	8.75	0.2240	7.71	0.130887	2.83
4	96	10	10	1.535	10.70	43.67	0.3814	15.98	0.300284	3.13
5	135	5	10	0.821	5.08	10.70	0.4146	6.08	0.093411	2.56
6	135	10	10.1	1.650	9.74	46.23	0.4873	12.22	0.330000	4.50
7	95	5	30	0.795	8.05	12.70	0.1960	8.34	0.111692	3.14

As presented in Table 2, the values of dispersion number approximately vary from 0.09 to 0.68, which are much higher than 0.01, thus the flow of particles within the dryer can be considered as a dispersed plug flow with large deviation from ideal plug flow. The values of dispersion coefficient range from 2.60×10^{-4} to 4.50×10^{-4} m^2/s over the investigated conditions. Typical results of dispersion coefficient for material somewhat smaller than rough rice are on the order of 1×10^{-4} to 20×10^{-4} m^2/s [7].

Figure 2 shows the cumulative residence time distribution F(θ) against θ, for all the RTD measurement experiments. As shown, the flow pattern for all the experimental conditions is between plug flow and well-mixed flow.

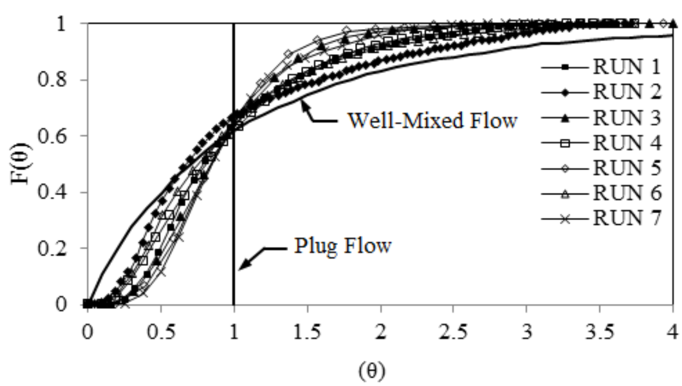

Figure 2. The Cumulative Residence Time Distribution Curves (Runs 1-7)

The mean values of relative error of the Reay's equation in predicting the dispersion coefficient are 16.52% and 42.81% and those of the Nilsson's equation are 102.06% and 55.51%, respectively, for weir heights of 5 and 10 cm. However, none of these equations is found to predict the dispersion coefficient for both weir heights reasonably well, the Reay's equation can be employed to predict the dispersion coefficient for weir height of 5 cm.

3.2 Effect of Solid Feed Rate on RTD

The effects of solid feed rate on RTD for weir heights of 5 and 10 cm are shown in Figures 3 and 4, respectively.

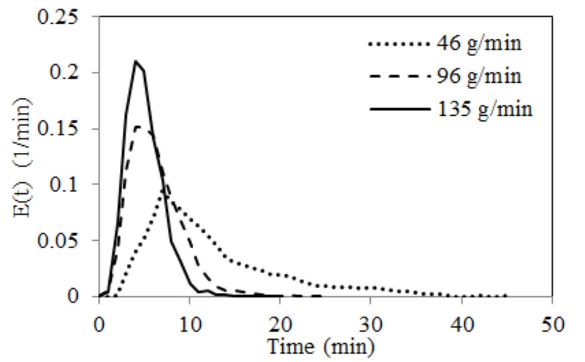

Figure 3. Effect of Solid Feed Rate on RTD (H = 5 cm)

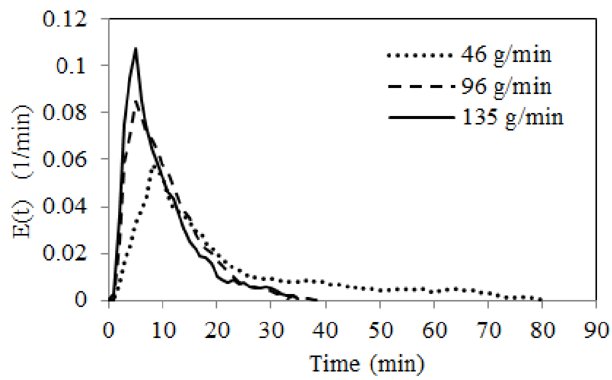

Figure 4. Effect of Solid Feed Rate on RTD (H = 10 cm)

The RTD curves exhibit asymmetry with generally steep gradient up to the maximum, followed by prolonged tail afterwards. As seen in Figures 3-4 and Table 2, solid feed rate has a significant effect on RTD characteristics. For a given weir height, the RTD curve shifts to the left with an increase in solid feed rate, thus the mean residence time is decreased. For example, the mean residence time decrease 59.29% and 54.80% when the solid feed rate increased from 46 to 135 g/min for 5 and 10 cm weir heights, respectively. The decrease of the mean residence time as solid feed rate increases at the fixed weir height considering the increased solid holdup is owing to the increased axial solid flow velocity in the dryer which in turn leads to the lower exit time of material from the dryer.

Also the increase in solid feed rate decreases the spread of the RTD curve about the mean. For example, increasing solid feed rate from 46 to 135 g/min, decreases the variance of distribution by 80.64% and 84.72% for 5 and 10 cm weir heights, respectively.

The increase of solid feed rate by 193% (from 46 to 135 g/min), increases the solid holdup by 19.8% and 24% for weir heights of 5 and 10 cm, respectively. The solid holdup does not increase proportionally with the increase of solid feed rate. Satija and Zucker [8] and Han, et. al., [9] have reported similar results about the effects of solid feed rate on the mean residence time, the variance of distribution, and the solid holdup in a pilot scale vibrated plug flow fluid bed dryer.

3.3 Effect of Solid Feed Rate on Cumulative Residence Time Distribution

The effects of solid feed rate on the F(t) for weir heights of 5 and 10 cm are shown in Figures 5 and 6, respectively.

As shown in Table 2, at weir height of 5 cm, increasing the solid feed rate reduces the dispersion number, but at weir height of 10 cm, the dispersion number decreases with solid feed rate up to 135 g/min and then it increases with an increase in solid feed rate. It should be mentioned that this increase is not significant.

At weir height of 5 cm and solid feed rate of 46 g/min, F(t) shows that around 70% of the particles are removed after spending between 2-14 min in the dryer, while the remaining 30% are removed after a further 31 min. At weir height of 5 cm and solid feed rate of 96 g/min, around 88% of the particles spend between 2-10 min in the dryer, while the remaining particles are removed after a further 15 min. At weir heights of 5 cm and solid feed rate of 135 g/min, almost 90% of the particles have a residence time in a relatively narrow range of 2-8 min (Figure 5).

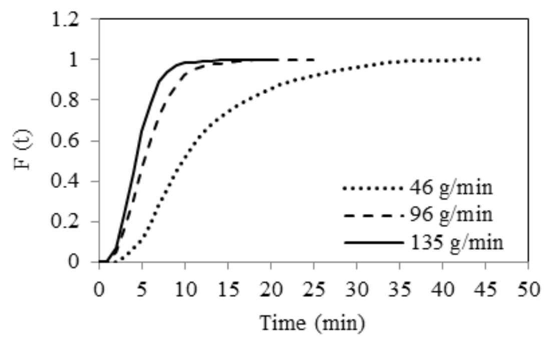

Figure 5. Effect of Solid Feed Rate on F(t) (H = 5 cm)

As shown in Figure 6, at weir height of 10 cm and solid feed rate of 46 g/min, about 64% of the particles leave the bed after spending between 1-20 min inside the bed, while the remaining particles are removed after a further 60 min. At weir height of 10 cm, the shape of F(t) curves at solid feed rates of 96 and 135 g/min are relatively similar. In these conditions, about 85% of the particles have a residence time in the range of 2-17 min and the remaining particles have a residence time of 18-35 min.

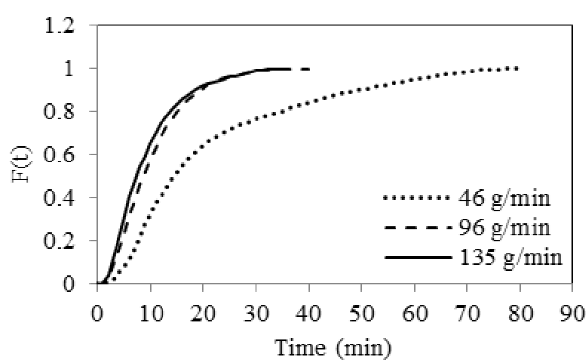

Figure 6. Effect of Solid Feed Rate on F(t) (H = 10 cm)

3.4 Effect of Weir Height on RTD

The effect of weir height on the RTD at solid feed rate of 96 g/min is shown in Figure 7. The increase in weir height from 5 to 10 cm causes the RTD curve to shift right thus the mean residence time and the variance of distribution are increased.

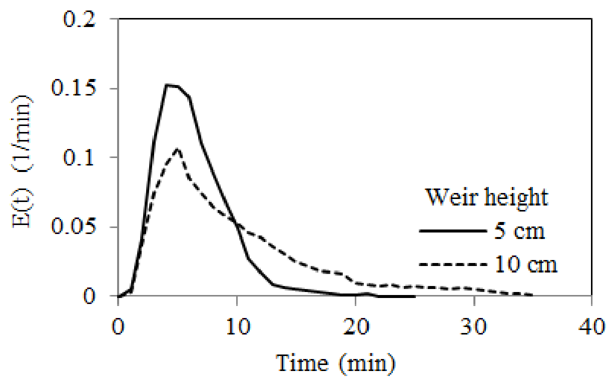

Figure 7. Effect of Weir Height on RTD (F=96 g/min)

As shown in Table 2, with a 100% increase in weir height, the mean residence time increased 171.68%, 71.20%, and 91.73% for solid feed rates of 46, 96 and 135 g/min, respectively. It can be mentioned that for a given solid feed rate, increasing the weir height increases the solid holdup, thus the mean residence time of the particles in the dryer is increased.

By increasing the weir height from 5 to 10 cm, the variance of distribution increased 447.34%, 399.08%, and 332.06%, the dispersion number increased 194.23%, 129.42%, and 253.28%, and the dispersion coefficient increased 51.54%, 10.60%, and 75.78% for solid feed rate of 46, 96, and 135 g/min, respectively (Table 2). The reason for this is that increasing weir height at a constant solid feed rate results in more bed height, and thus more axial mixing, which lead to increased variance of distribution, dispersion number, and dispersion coefficient. Nilsson [6] also reported that a deeper bed has a larger dispersion coefficient and vice versa.

As expected, there is an increase in solid holdup about 94.16%, 107.43%, and 100.97% with a 100% increase in weir height for solid feed rates of 46, 96 and 135 g/min, respectively. To some extent, it can be concluded that the solid holdup is linearly related to the weir height in a plug flow fluid bed dryer.

3.5 Effect of Inlet Solid Moisture Content on RTD

As shown in Table 2 and Figure 8, at a given solid feed rate and weir height, the mean residence time increased from 6.25 to 8.05 min by increasing the inlet solid moisture content from 10 to 30% d.b. Finzer, et. al., [13] and Renaud, et. al., [14] investigated the vibrated drying of coffee and the rotary drying of sand, respectively, and reported that the mean residence time during the drying tends to decrease as moisture content decreases.

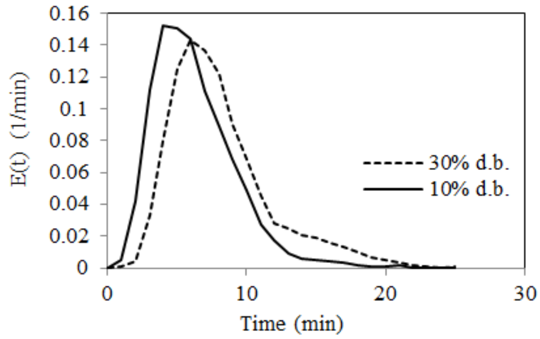

Figure 8. Effect of Inlet Solid Moisture Content on RTD (F = 96 g/min, H = 5 cm)

As shown in Table 2, the increment of inlet solid moisture content from 10 to 30% d.b. at solid feed rate of 96 g/min and weir height of 5 cm, leads to a reduction in dispersion number and dispersion coefficient from 0.130887 to 0.111692 (about 14.7% decrease) and from 2.83×10^{-4} to 3.14×10^{-4} m^2/s (about 10.9% decrease), respectively. Higher inlet solid moisture content may result in somewhat sticky particles, thus some closer behavior to the plug flow, resulting in less dispersion coefficient.

Nilsson [6] in order to consider the decreasing effect of solid moisture content on dispersion coefficient in a vibrated plug flow fluid bed dryer assumed that the dispersion coefficient increases linearly along the length of the dryer.

3.6 Comparison of Mean and Linear Residence Times

In ideal plug flow pattern, there is no stagnant region in the bed, thus the values of mean residence time (τ_R) calculated based on RTD and the linear residence time (τ_H) calculated based on solid holdup and solid feed rate are the same. In situation, where there exists a dead region in the dryer, the tracer response is occurred early (mean of RTD curve is somewhat early) and RTD curve has a long tail, thus the values of τ_R and τ_H are different. The values and the parity plot of τ_R and τ_H are given in Table 2 and Figure 9, respectively. As seen, the values of τ_R and τ_H are in good agreement with each other; however the values of τ_R are somewhat lower than those of τ_H. Low values of this difference could be due to the existence of small stagnant or dead region within the bed. In this study, the existence of the stagnant region within the bed is probable near the exit weir.

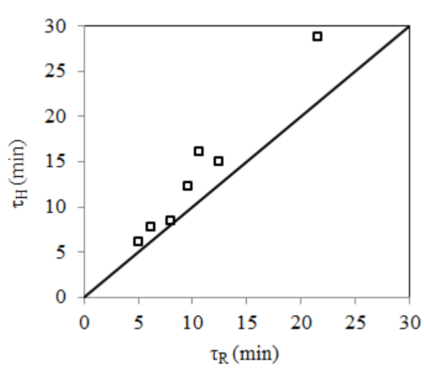

Figure 9. Comparison of Mean Residence Time and Linear Residence Time

4. Conclusions

Knowledge of RTD provides useful information for identification of flow pattern in a plug flow fluid bed dryer. The results obtained are beneficial for the design, modeling, control, and scale-up of the process and the unit operation. The values of dispersion number and dispersion coefficient, respectively, range from 0.09 to 0.68 and from 2.60×10^{-4} to 4.50×10^{-4} m^2/s, over the investigated conditions. Thus the flow of particles in the dryer can be considered as a dispersed plug flow with large deviation from ideal plug flow. The increase of solid feed rate results in a decrease in mean residence time and increase in axial mixing. The flow in the plug flow fluid bed dryer approaches plug flow as the weir height is decreased, whereas a decrease in the inlet solid moisture content results in a well-mixed flow. Low values of difference between mean residence time and linear residence time indicate less stagnancy of material in the dryer.

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