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Developing and Assessing Performance of a Laboratory-Scale Fluidized Bed Dryer

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Authors' contributions

This work was carried out in collaboration between both authors. Both authors read and approved the final manuscript.

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ABSTRACT

A laboratory-scale batch fluidized bed dryer with 75 mm bed diameter was designed, fabricated and evaluated to study the hydrodynamics of river sand as well as the drying of cassava mash and bitter kola particulates. The hydrodynamics properties such as minimum fluidization velocity, effect of bed height and pressure drop across the bed, effect of particle size and density on minimum fluidization velocity and stability of the bed column of river sand were studied. Drying of cassava mash to edible garri was carried out at in fluidized bed dryer at controlled temperatures. Drying characteristics of the laboratory fluidized bed dryer was compared with the laboratory WiseVen oven (model number: WOF - 105) using bitter kola particulate material sample. The experimental results from laboratory-scale fluidized bed dryer showed that the minimum fluidization velocity increases as material density increases and the value of the minimum fluidization velocity obtained from the fluidized bed gave good agreement with other empirical correlation such as Kozeny-Carman Equation. The fluidized bed dryer showed high rates of moisture removal over the conventional oven with the ratio of 1:29 under the same operating conditions. Drying of cassava mash to edible garri was achieved at lower drying temperatures of $83\,℃ \pm 3\,℃$ at 55 minutes when compared to conventional frying

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temperatures of cassava mash to edible at 180 – 200 ℃ thereby saving energy cost. Hence, this fluidized bed dryer is recommended for use in demonstrating hydrodynamics and drying of particulate materials in the laboratory.

Keywords: Fluidized bed dryer design; hydrodynamics; heat transfer; minimum fluidization velocity; drying.

1. INTRODUCTION

Fluidization is a process whereby fine solids are transformed into a fluid-like state in the fluidized bed through contact with a gas [1,2]. Fluidized bed operations are usually carried out within a given flow regime [3]. Due to favourable heat and mass transfer characteristics of fluidization, fluidized beds are primarily used for gas-solid contacting processes [4]. According to [5] and [6], moving particles play an important role in the wall-to-bed heat transfer processes of which wall-to-bed heat transfer processes is higher than those in single-phase gas flow as well as those in fixed beds. Fluidized bed has advantages in industrial applications in areas involving heterogeneous heat transfer, drying, mass transfer and chemical reactor due to the excellent contacting ability between the solid and fluid phase [7]. Souraki, et al [8] showed that mixing and heat transfer processes are very rapid and the exit gas usually saturated with vapour for any allowable fluidization velocity.

Studies showed that drying process is greatly affected by the internal heat transfer whereas the gas pressure distribution effect is insignificant [9]. Bouhadda et al [10] studied the heating of gas in fluidized bed of sand particles irradiated indirectly by concentrated solar energy and pointed out that the heating process is strongly affected by the gas fluidizing velocity and the wind speed. Zhang et al [11] described convective and radiative heat transfer gas-solid in fluidized bed whereas [12] had shown the effect of radiative heat transfer contribution to total heat transfer of about 13% and 18% for at high operating temperature of 400 °C and 600 °C respectively in the gas-solid in fluidized bed. Studies showed that there are three distinct mechanisms of heat transfer in fluidized beds namely (i) fluid-to-particle, (ii) particle-to-fluid and (iii) wall-to-bed [13,14].

Heat transfer in gas-fluidized beds may be described as a complex mutual interactions of fluid mechanics, particle dynamics and the thermal properties of the involved media [15]. Good knowledge of the heat-transfer characteristics between the fluidization moving particles and the dense bed are important for the design of units utilizing the fluidized bed system [16]. Zhang et al [17] showed that heat transfer characteristics are relevant at transition between packed and fluidized bed operations. Studies showed that fluidized bed is the most useful method for drying particles or granules in pharmaceutical and food industries [18,19] due to increase in surface area of the particles [20]. Thermal efficiency of the fluidized bed dryers is most important item for variety of drying applications [21].

According to Sundaresan [22], the particle-phase pressure arising from flow-induced velocity fluctuations decreased with increased in concentration of particles. Ergun equation showed that the ratio of pressure gradient to superficial fluid velocity in a column is a linear function of fluid mass flow rate, and the constants of this linear relationship are particle specific surface, fractional void volume, and fluid viscosity [23]. Venkiteswaran et al [24] showed that in fluidized bed processes, bed pressure drop is crucial as it determines the pumping power required for fluidization. Sasic et al [25] used pressure signal to investigate the fluid-dynamic behaviour of gas–solid fluidized beds in comparison to those obtained from modelling or experimentation and [26] used pressure fluctuations to investigate the effect of particles distribution in gas–solid fluidized beds. Fluidization processes have wide application in many industries ranging from heavy chemicals, mining, food, fine chemicals, petroleum, and pharmaceutical industries [27]. For drying of powders in the 50 to 2000 µm range, fluidized beds compete successfully with other dryer, such as: rotary, tunnel, conveyor, and continuous tray [28,29].

Studies showed that fluidized bed dryers are classified based on operating pressures, processing mode, fluidizing gas flow and fluidized materials. According to Mujumdar et al [28], the advantages of fluidized bed dryer include: ease of control, high drying rates, smaller flow area, higher thermal efficiency, lower capital and maintenance costs when compared to rotatory dryers. However, Mujumdar et al [28] pointed out some limitations such as: not generally recommended for drying materials when organic solvents need to be removed, higher power consumption needed to suspend the entire bed in gas phase, increase in gas handling, high potential of attrition due to granulation or agglomeration and potential of de-fluidization for feed that is too wet.

To overcome some of the limitations or problems encountered with conventional fluidized bed dryer during drying processes, other modified fluidized bed dryers were developed for used by [30]. They are: hybrid fluidized bed dryers, pulsating fluidized bed dryers, fluidized bed dryer with immersed heat exchanger, mechanically assisted fluidized bed dryer, vibrated fluidized bed dryer, agitated fluidized bed dryer/swirl fluidizers, fluidized bed dryers of inert particles, spouted bed dryer, recirculating fluidized bed dryer, jetting fluidized bed dryer, superheated steam fluidized bed dryer, fluidized bed freeze dryer and heat pump fluidized dryers

The drying application of fluidization technique for particulate materials in industry dated as far back 1940s [31]. According to Oluwaleye et al [32], fluidized bed drying has the advantage of high intensity of drying and high thermal efficiency with controllable temperature due to high rates of heat and mass transfer which reduces the drying time in the fluidized bed dryer. It is convenient to dry heat sensitive food materials in fluidized bed dryer as it prevent them from overheating due to its mixing characteristics [33]. The drying process with fluidized bed drying reduces the drying time in the drying when compared with conventional oven by ten to twenty times [34]. According to Ali et al [35], gelatinization temperatures during garri processing are within (70 - 90 ℃), and frying temperatures are within (180 - 200 ℃) using traditional methods. Obadina et al [36] used rotary dryer to dry cassava mash to edible garri. Ismail et al [37] developed a fluidized bed dryer with centrifugal blower incorporated with 9 kW air heater to fry 2 kg of garri at drying temperature of 120 -150 ℃ within 18 - 24 minutes.

2. METHODOLOGY

2.1 Mathematical Modelling of Fluidized Bed Unit

According to Perry and Green [38], the major parts of a fluidized-bed system include:

i. Fluidization vessel which comprises (a) fluidized-bed portion, (b) disengaging space or freeboard, and (c) gas distributor
ii. Solids feeder or flow control
iii. Solids discharge
iv. Dust separator for the exit gases
v. Instrumentation
vi. Gas supply

2.1.1 Minimum fluidization velocity

Minimum fluidization velocity (U_{mf}) which could be determined using the Ergun equation, Kozeny-Carman correlation, voidage method, correlation of Wen and Yu, the heat transfer method, or the pressure drop method [39,40]

2.1.2 Flow regime

Flow regime which indicates whether the flow is laminar or turbulent is defined by Equation (1).

$$Re_{mf} = \frac{U_{mf} d_p \rho_g}{\mu_g} \tag{1}$$

2.1.3 The void fraction

The void fraction or porocity (ε) depends on the material, shape, and size of the particles. For nearly spherical particles, [41], suggested that **ε** laid in the range 0.4 - 0.45 and therefore increasing with particle size and could be calculated from Equation (2), where bulk density, ρ_b, and ρ_s is the particle density.

$$\varepsilon = 1 - \frac{\rho_b}{\rho_s} \tag{2}$$

2.1.4 Maximum or terminal velocity

Maximum or terminal velocity or the settling velocity is given in Equation (3) for low Reynolds number, where Re < 500; and Equation (4) high Reynolds number, where 500<Re<20000.

$$U_t = \frac{d_p^2(\rho_s - \rho_g)g}{18\mu_g} \tag{3}$$

$$U_t = \frac{1.75\sqrt{d_p(\rho_s - \rho_g)}}{\rho_s} \tag{4}$$

2.1.5 Bed sizing

According to Perry et al [38], the cross-sectional area of fluidized bed reactor is determined by the volumetric flow of gas and the allowable or required fluidizing velocity of the gas at operating conditions. The maximum flow is generally determined by the carry-over or entrainment of solids, and this is related to the dimensions of the disengaging space (cross-sectional area and height). Also, Perry et al [38] showed that bed heights are not less than 0.3 m (12 in) or more than 16 m (50 ft) for fluidized bed dryers. Based on cost evaluation, space, and TDH chart in Fig. 1 [42], a 7.5 cm internal diameter fluidized bed was considered for this work. Determination of the bed height involved additionally the following hydrodynamic parameters:

i. Minimum fluidization velocity was determined using the expression Kozeny-Carman [43] in Equation (5).

$$U_{mf} = \frac{d_p^2(\rho_s - \rho_g) g\, \Phi_s^2}{150 \mu_g} \left(\frac{\varepsilon_{mf}^3}{1-\varepsilon_{mf}}\right) \quad (5)$$

ii. Terminal or maximum velocity was calculated from Stokes law using in Equation (6).

$$U_t = \frac{d_p^2(\rho_{s-}\rho_g) g}{18 \mu_g} \quad (6)$$

iii. The superficial velocity of the gas during the operation considering the relation between the expanded and minimum heights of the fluidized bed H/ H_{mf} , according to [44] is given Equation (7).

$$\frac{H}{H_{mf}} = 1 + \frac{10.978 \,.(U - U_{mf})^{0.738} \,.\, \rho_p^{0.376} \,.\, d_p^{1.006}}{U_{mf}^{0.937} \,.\, \rho_g^{0.126}} \quad (7)$$

For the bubbling fluidized bed the restriction suggested in Equation (8) was used [5]:

$$1.2 < \frac{H}{H_{mf}} < 1.4 \quad (8)$$

For this design, a value of 1.3 was selected for H/ $\boldsymbol{H_{mf}}$, and Equation (7) was solved to determine the value of fluidization velocity *(U - U_{mf})*

iv. Transport Disengaging Height, TDH, was extrapolated from the graph of TDH vs. *(U - U_{mf})* in Fig. 1.

v. Overall height Ht, of the fluidized bed chamber was established by the expression shown in Equation (9) [5]:

$$H_t = TDH + H \quad (9)$$

vi. The maximum expanded height of the bed H, was assumed as 0.15 m, being twice the internal diameter of fluidized bed, with the purpose of diminishing the slugging phenomena. Yang et al [45] considered that the slugging regime appears in beds where the bed height (H) over the bed diameter (D) is larger than two (2). This requirement ensures that bubbles have enough time to coalesce in bigger bubbles called slugs, when the bubbles grow to two-third of the bed diameter the system enters to a slugging regime.

2.1.6 Wind box

The gas stream nozzle was made to enter the gas chamber vertically through the bottom. To prevent mal-distribution, the nozzle should be placed at a distance H_w below the distributor plate [46], where H_w is given by:

$$H_w = 3 \times (D_w - D_{noz}) \quad \textit{for} \;\; D_{noz} > \frac{D_w}{36} \quad (10)$$

$$\text{Or, } H_w = 100 \times D_{noz} \quad \textit{for} \;\; D_{noz} < \frac{D_w}{36} \quad (11)$$

Where D_w is the bed diameter, D_{noz} is the gas nozzle diameter or the gas entry pipe, and H_w is the distance of the nozzle from the distributor plate.

2.1.7 Air distributor plate

Perforated distributor plate design was adopted for this work considering the following design criteria:

i. Karri et al [47] showed that $\Delta P_{gird} \geq 0.3\, \Delta P_{bed}$ for upwardly and laterally directed flow, and under no circumstances should the pressure drop across a large-scale commercial grid be less than 2500 Pa.

Also, the gas velocity through the grid hole, *(U_h)*, is related to ΔP_{gird} in orifice as shown in Equation (12)

$$\Delta P_{gird} = \frac{U_h^2 \rho_g}{2C^2} \quad (12)$$

$$\text{And} \quad U_h = C\sqrt{\frac{2\,\Delta P gird}{\rho_g}} \quad (13)$$

U_h is the velocity in hole at inlet condition, ρ_g is the density of the fluid, ΔP_{grid} is the pressure drop, and C is the orifice coefficient or constant, dimensionless (typically 0.8 for gas distributors)

ii. Number of grid holes required is related to volumetric flow rate, Q

$$Q = ANU_h \quad (14)$$

Where Q is the volumetric flow rate in the holes (m/s), N is the number of holes, and A is cross-sectional area of the bed (m^2).

iii. The hole density, N_d, is defined as the number of grid holes required per unit area.

$$N_d = \frac{N}{A} \quad (15)$$

This work considered grid holes pitch configuration in a triangular arrangement as given by Equation (16)

$$L_h = \frac{1}{\sqrt{Nd \sin 60}} \quad (16)$$

2.2 Particulate Material Selection and Preparation

For this study, particulate materials were selected using the four criteria: (i) fluidization behavior, (ii) size range, (iii) density, and (iv) aspect ratio [48] while cassava mash and bitter kola were prepared using local grating method.

2.3 Experimental Set-up

The schematic of experimental set-up from [49] shown in Fig. 2 consist of air compression system (COM 101 and CAM 101), air shut off vale (ASV 101), pressure control valve (PCV 101), flow temperature indicators (TI), helical coil heat exchanger (HCHE), fluidized bed vessel (FBV), differential pressure manometer (DP), flow meter (rotameter), pressure gauge (PG), temperature controller (TC) and level indicators (LI). In this system hot air for fluidization drying was produced from heat transfer in HCHE.

2.3.1 Hydrodynamics studies in fluidized bed

Hydrodynamics experiments were conducted at ambient conditions. Particulates materials were poured into the bed up to the desired bed height/weight of particles. Hence, the following procedures were considered:

i. Measurement of the bed heights directly using a scale attached along the height of column.
ii. Switch on the mains supply.
iii. Start the compressor to compress the air above 100 psi.
iv. Open the valve upstream of the air flow meter and adjust the flow rate to achieve good fluidization
v. Measure air flow rate and pressure drop from the instruments.
vi. Plot pressure drop against superficial air velocity to determine minimum fluidization velocity according to [5].

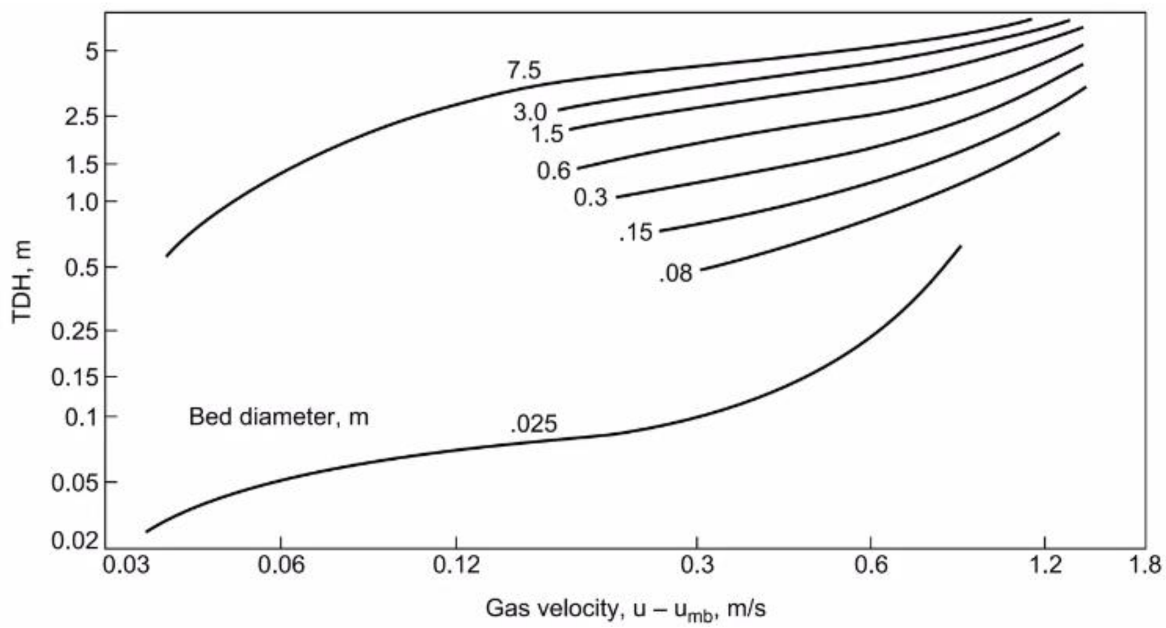

Fig. 1. The graph of TDH vs. *(U - U_{mf})*
Source: [42]

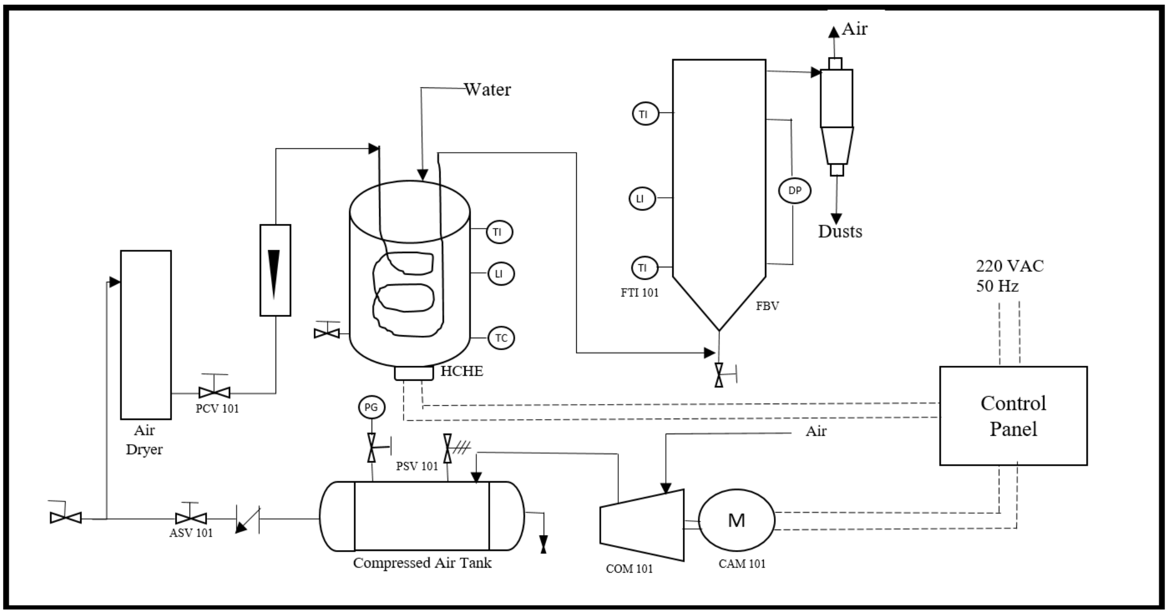

Fig. 2. Experimental set-up
Source: [49]

2.3.2 Drying of cassava mash to Garri in fluidized bed dryer

The drying of cassava mash to edible garri was carried out in the fluidized bed dryer with the ratio of the bed height to bed diameter (H/D) of 1.8. The initial mass of cassava mash was 224.76 g with the mean particle size of 834 µm. The drying temperature was 50 ± 3 ^{0}C at initial stage for 15 minutes to avoid gelatinous formation. During this drying period, fluidization of the bed material was supported by vibrating the bed. This was due to the damp and sticky nature of the cassava mash particulates material. From 20 minutes of the drying period, the particles where loose and the mixing was improved up to 35 minutes and the drying temperature was set at 70 ± 3 ^{0}C. Thereafter, fluidizations were adequate till the end of drying time of 55 minutes in which the drying temperature was controlled at 83 ± 3 ^{0}C. The mass of the cassava mash sample during drying was recorded at intervals 5 minutes and the corresponding moisture content were determined.

3. RESULTS AND DISCUSSION

3.1 Design Parameters for Fluidized Bed Column, Plenum and Distributor Plate

Design parameters of the fluidized bed column as shown on Table 1

Design parameters for plenum and distributor plate as shown on Table 2

Table 1. Design parameters of the fluidized bed column

S/N	Parameters	Value
1	Minimum Fluidization Velocity U_{mf},	0.12 m/s
2	Terminal Velocity U_t	9.6 m/s
3	Calculated Value of (U- U_{mf})	0.49 m/s
4	Design value of (U- U_{mf})	0.5 m/s
5	Bed height, L_B	0.15 m
6	Transport Disengaging Height, TDH	0.8 m
7	Bed height, L_B	0.15 m
8	Total Bed Height H_t	0.95 m

Table 2. Design parameters for plenum and distributor plate

S/N	Parameters	Value
1	Fluidized Bed Diameter D_w	75 mm
2	Nozzle Diameter D_{noz}	20 mm
3	Nozzle distance from the distributor plate H_w	0.165 m
4	Orifice Coefficient C	0.78
5	Grid Plate Thickness t	1.5 mm
6	Grid Hole Diameter d_h	4 mm
7	Pressure drop across the distributor plate ΔP_{grid}	2500 Pa
8	Volumetric gas flow Q	2.2×10^{-3} m3/s
9	Number of grid holes N	3.3
10	Hole density, N_d	750/m^2
11	Holes pitch length, L_h	39 mm
12	Grid hole velocity U_h	52.58 m/s

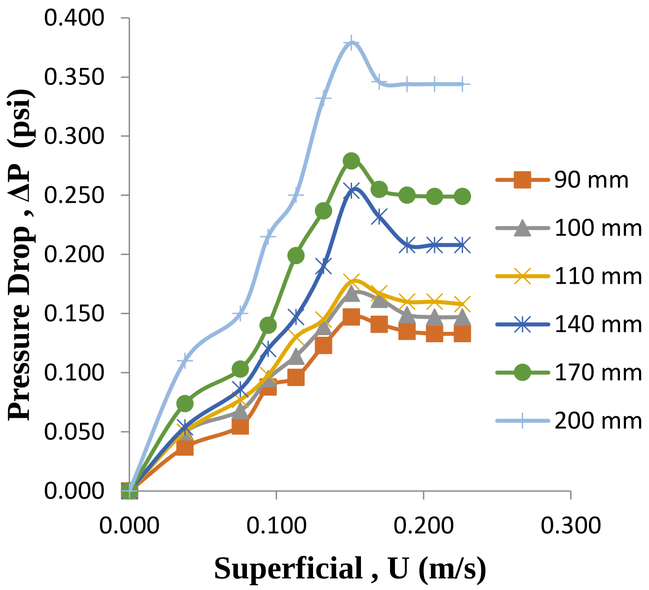

Fig. 3. The relationship of change in bed pressure drop and superficial Velocity

3.2 Experimental Determination of Minimum Fluidization Velocity

The relationship of change in bed pressure drop and superficial velocity to determine minimum fluidization velocity of river sand particles with mean diameter, d_p, of 521 µm using different bed heights were carried out and the graphical representation as shown in Fig. 3, indicated the minimum fluidization velocity of 0.15 m/s.

3.3 Comparison of Experimental and Predicted Minimum Fluidization Velocity

The experimental result and the empirical correlations as proposed by Kozeny-Carman [43], [50 - 52], were analysed in Table 3. There was a good agreement of the experimental value with prediction of Kozeny-Carman [44]. According to Baeyens et al [52], some forms of predictive correlation entailed relatively large error of 30-40% and is only valid for limited conditions as shown in Ergun [50] and [51]. This errors are due to these correlations relying on a small number of specific experiments and in some cases do not take into consideration the effect of particle sphericity [52].

3.4 Effect of Bed Height and Pressure Drop across the Bed on the U_{mf}

The readings of bed pressure drops were recorded by increasing superficial velocity until the packed bed reached its minimum fluidization velocity. These sets of experiments were carried

Table 3. Comparison of experimental and predicted values for minimum fluidization velocity

Author	Experimental /Empirical Correlation	Errors Based on Experiment Value			
		Values	**Absolute Error**	**% error**	**Relative error**
Experiment	Experiment	0.1500	0.0000	0.000	0.0000
Kozeny-Carman [43]	$U_{mf} = \frac{d_p^2(\rho_s-\rho_g)g\,\Phi_s^2}{150\mu_g}\left(\frac{\varepsilon_{mf}^3}{1-\varepsilon_{mf}}\right)$	0.1530	0.0030	0.300	0.0030
Leva [59]	$U_{mf} = 9.23\text{x}10^{-3} d_p{}^{1.811}\left(\frac{\rho_g}{\mu_g}\right)^{0.88}\left(\frac{\rho_s}{\rho_g}\right)^{0.94}$	0.1800	0.0300	20.000	0.2000
Ergun [50]	$U_{mf} = \frac{(\rho_s-\rho_g)g}{1650\mu_g}$	0.2180	0.068	45.300	0.4530
Sidorenko [51]	$U_{mf} = \frac{0.0093 d_p^{1.82}(\rho_s-\rho_g)^{0.94}}{\mu^{0.88}\rho_g^{0.06}}$	0.2200	0.0700	46.700	0.4700
Baeyens [52]	$U_{mf} = \frac{0.000701 d_p^2(\rho_s-\ \rho_g)g}{\mu}$	0.2500	0.1000	66.700	4.4000

out with different bed heights and different particle sizes distributions as shown on Fig. 4 and 5. The figures represented the relationship between pressure drops with different bed height for 400 µm, and 500 µm. From the figures, it showed that increase in bed height in fluidized bed led to increase in the pressure drop across the packed bed and these results corroborate data presented by [53,54].

3.5 Effect of Particle Size and Density on Minimum Fluidization Velocity

The influence of particle sizes of the a particular sand sample with mean particle sizes of 400 µm, 500 µm, 521 µm with densities of 2357 kg/m^3, 2362 kg.m^3 and 2500 kg/m^3 respectively were analysed using the same bed height to bed diameter ratio of 1.5. Graphical representation indicated that particle size distribution influences the minimum fluidization velocity as shown in Fig. 5. Also, a denser material required a higher superficial gas velocity to start fluidization. Therefore, the minimum fluidization velocity increases as material density increases. These experimental results also corroborate data presented by [55].

3.6 Stability of the column

For better fluidization, the ratio of bed height to diameter of bed (H/D) is usually maintained between 1 and 2 [56] and [44]. This is with the purpose of diminishing the slugging phenomena, which is associated with pressure fluctuation and vibration of the fluidized bed vessel (FBV). The maximum expanded height of the bed according to the design was 150 mm being twice the bed height. The evaluation of the FBV stability was tested with several bed height (H), to bed diameter (D), (H/D ratios) from 1 to 2.6. It was observed that from the H/D of 2.2 and above, there was slugging in the bed which was accompanied by little vibration during fluidization. This phenomenon corroborates what was presented by [44].

3.7 Drying of Cassava Mash to Garri in Fluidized Bed Dryer

The time taken for drying of cassava mash to edible garri was 55 minutes at controlled drying temperature of 83 ± 3 ^{0}C against 120 – 150 ℃ [37] and 180 – 200 ℃ [35] by other conventional fluidized bed dryer and traditional methods respectively. The relationship of moisture content and time for garri particulate material as shown in Fig. 7 showed a steady state drying characteristics curve which corroborate the drying curve as presented by [58].

3.8 Comparing Drying of Bitter Kola in WiseVen Oven and Fluidized Bed Dryer

The base line moisture content of the sample was determined by drying 5 g of bitter kola particulate in oven at temperature of 105 ℃ until there was no change in mass and the moisture content removed was 2.56 g. Equal mass of bitter kola particulates each 170 g were dried in

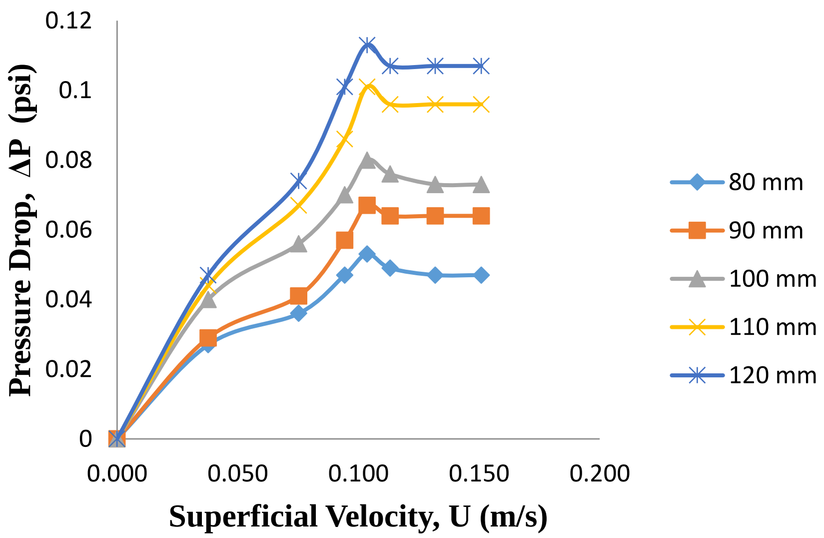

Fig. 4. Relationship of bed height with pressure drop on U_{mf} for 400 µm sand particle

Wise Ven oven and fluidized bed dryer at 50 ℃ for 60 minutes. The mass of moisture content removed from the oven studied was 2 g and that of fluidized bed dryer was 56.4 g. The results showed that 1.02 g of moisture was removed from the oven and that of fluidized bed dryer was 28.88 g representing the ratio of 1:29. The results corroborate data presented by [34]. Hence, the fluidized bed dryer has a high rates of moisture removal due to high heat and mass transfer rates over conventional oven considering the same operating conditions.

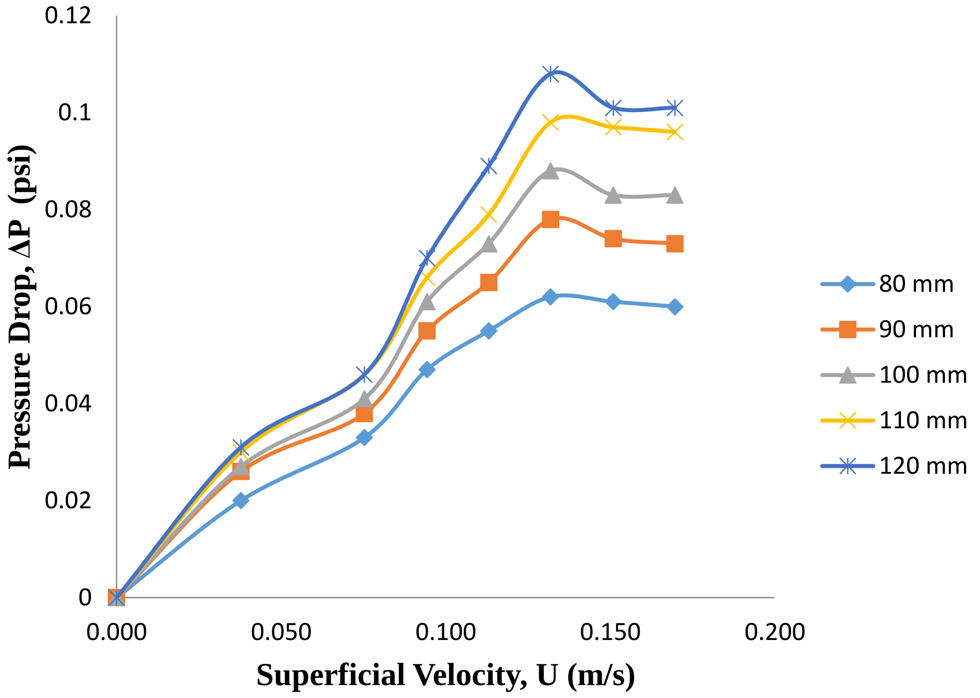

Fig. 5. Relationship of bed height with pressure drop on U_{mf} for 500 µm sand particle

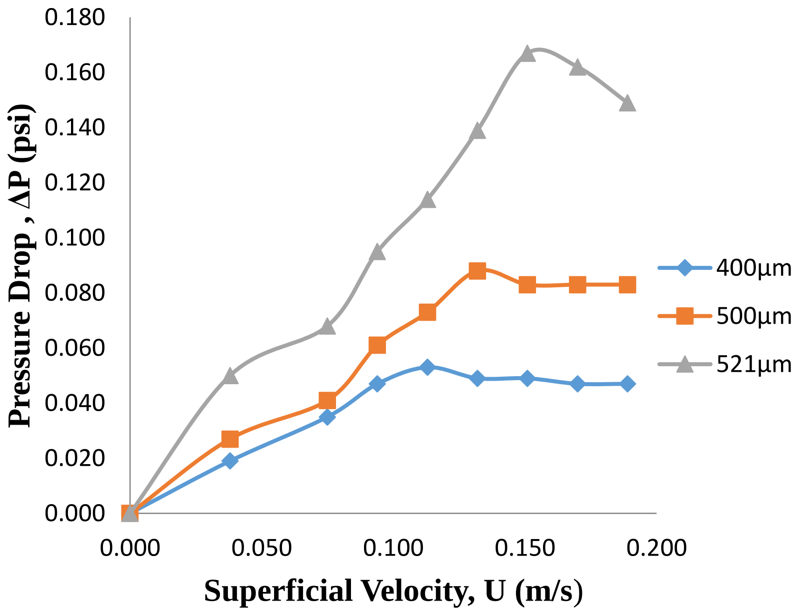

Fig. 6. Influence of particle size distribution and density on minimum fluidization velocity

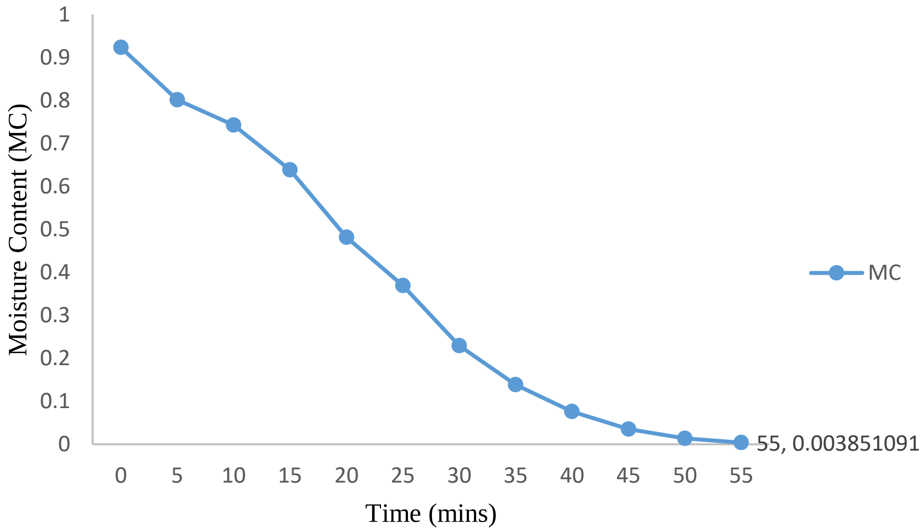

Fig. 7. Moisture content as a function of time for garri particulates material

4. CONCLUSION

Laboratory-scale fluidized bed dryer was designed, fabricated for fluidization hydrodynamics and fluidization drying. From the performance evaluation of the fluidized bed dryer, the following conclusions were drawn from the study:

i. The design was efficient for laboratory applications.

ii. The unit was designed for non-corrosive and Galdart group A and B materials only.

iii. Excellent fluidization was achieved with particulate materials whose aspect ratio was tending to unity.

iv. Drying rate was effective in the ratio of 1:29 over conventional oven

v. The design was made for batch operations.

vi. The operational stability of this equipment with excellent fluidization of particles in the bed occurred between the H/D ratios of 1.5 – 2.

vii. Fabricated laboratory scale FDB offers excellent drying performance. Interestingly, cassava mash were dried to edible garri within 55 minutes at maximum temperature of 83 ± 3 ^{0}C against 120 – 150 ℃ and 180 – 200 ℃ by other conventional fluidized bed and traditional methods respectively.

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COMPETING INTERESTS

Authors have declared that no competing interests exist.

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