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Modeling and Simulation of Fluidized Bed Drying of Chickpea

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Abstract: A model for drying chickpea seeds in a batch fluidized bed has been dealt with in the present paper assuming that the bed is divided into a solid phase, an interstitial gas phase and a bubble phase with heat and mass transfer between the phases. The model considers further the bubble size, bubble rise velocity and bubble volume fraction in the rate equation for the constant rate and falling rate periods. The experimental set up consisting of an air heater and a fluidization column chamber with a provision of taking samples at a regular interval. The experimental runs for studying various effects of volumetric flow rate, initial moisture content of chickpea, its diameter, diffusivity have been compared with the simulated data found by running the listed MATLAB programme using the established working equations related with fluidized bed drying. Also the effect of varying the fluidized bed chamber-column has been studied using the MATLAB.

Keywords: Chickpea, Drying, Constant rate period, Falling rate period, Fluidized bed drying, Simulation, MATLAB.

1. INTRODUCTION

The objective of drying of agricultural product is long time storage without quality deterioration. Chickpea is a species of pulse. The physical properties of chickpea used in the present work include particle sizedp 9 mm-13 mm, particle densityρ_p1280 Kg/m^3, specific heat c_p 1980 J/KgK, with average moisture content 80 gm/Kg. Desi chickpea is available mostly in semi-arid tropical area of Indian sub-continent, whereas Kabulitype chickpea is available in temperate regions. Out of some 50 countries growing chickpea, India ranks top in terms of its production and consumption. The chemical properties of chickpea are comparable with wheat. The analysis of chickpea starch includes ash, moisture and dry matter on percentage basis as 0.5, 11.0, 88.5 respectively. The fundamental drying theory includes the moisture removal by convection process with the vapour pressure of the product to be dried is greater than that of ambient. Drying process may be of capillary flow being taken into consideration in the present work, liquid diffusion , i.e., liquid movement owing to moisture concentration difference, surface diffusion ,i.e., liquid movement because of moisture diffusion of pore spaces, vapour diffusion, i.e., vapour movement due to driving force of moisture concentration difference, thermal diffusion , i.e., vapour movement with the driving force of temperature difference and hydrodynamic flow , i.e., water and vapour movement by means of the driving force of total pressure difference.

There may be of various types of dryer including Solar, Impingement, Pneumatic, Flash, Spray, Infrared, Microwave Tray, Freeze, Rotary, Spouted, Centrifugal, Fluidized. Open air drying of chickpea needs some 20 days in summer time for complete drying at the cost of the product quality in terms of loss of nutrients. Keeping in view the drying as the energy-intensive and negative environmental impact, consideration of the indirect-type solar dryer reduces average moisture content to 95% requires an average of 35 hours. The chickpea seeds are recommended to be dried at a low temperature like 15^{0}C and relative humidity of 15% to maintain their quality and longevity. Because of various advantages including good mixing with better heat and mass transfer with higher thermal efficiency with closely controllable temperature in the fluidized bedas drying gas mixed intensively during its passing through the fluidized bed with shorter time required for drying the same mass, absence of hot spot, the bubbling fluidized bed technique is being widely used at a commercial scale for drying of grains in various industries like pharmaceutical, fertilizer and food. There may be having some disadvantages including high pressure drop, attrition of solid and non-uniform moisture content in the product output, but these disadvantages are predominated by more advantageous points. For small scale operation batch operation is preferred because of the homogeneity particularly for the heat sensitive material. Industrial fluidized bed dryer development for any particular application of chickpea is somewhat difficult because of uncertainty in proper scale-up and uniform product quality. This unpredictability is because of different behavior of bubbles and mixing regimes in fluidized bed dryers of different sizes. The fluidized bed drying may be cross flow type recommended for free or weekly bound moisture and the other type may be counter-current recommended for strongly bound moisture. However, it is reported that on operating cost basis, the multi-stage concurrent flow dryer is recommended in comparison with the cross flow dryer for drying the seeds like chickpea.

The modeling and simulation of a fluidized bed dryer needs the understanding of the complex combination of hydrodynamics and mass transfer equations which are quite essential for optimization of the operating fluidized bed dryers and design of improved and more efficient fluidized bed dryer. There is various research papers (1-

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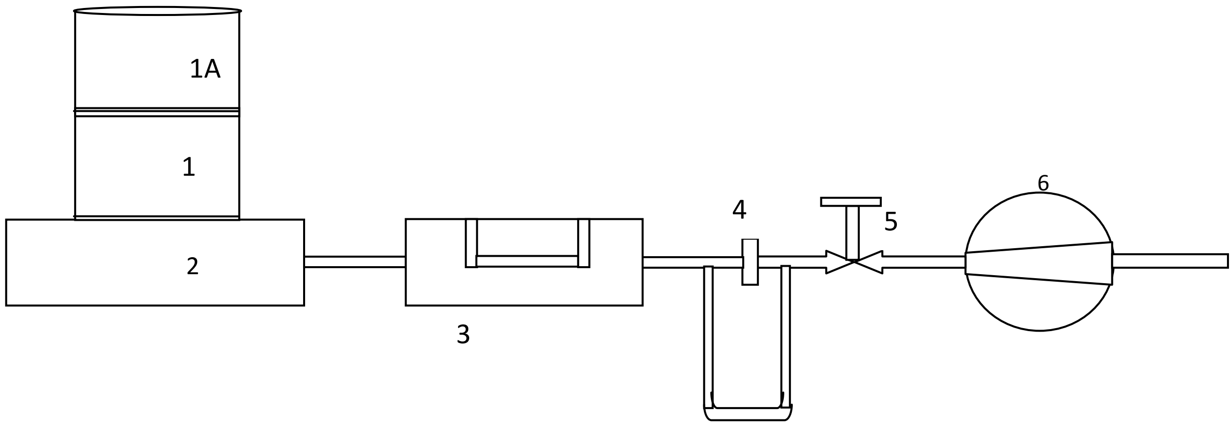
19) dealt with the various operating parameters controlling the fluidized bed drying.

Most of the agricultural products' drying take place in falling rate drying period. The present paper has ignored warming up drying rate period separately. The mode of moisture removal by drying is by internal diffusion and the temperature difference between the centre of the chickpea particle and the fluidizing medium like air. There are two distinct drying zones namely constant rate period and falling rate period. These two zones are differentiated by critical moisture content. Modeling of chickpea having the tendency of high resistance for moisture diffusion can be done by simple exponential time decay model like Newton, Page, Henderson and Pabis. The model dealt with and taken help of is based on differential equations and the fluidized bed dryer is assumed to have been divided horizontally and vertically into major and minor control volume.

A typical fluidized bed dryer set up for drying conventional material is shown as below as fig.1.

Figure1. Fluidized bed dryer set up for drying conventional material.

A laboratory-scale fluidized bed drying system was got fabricated and used as shown in the following figure 1 showing the various simple parts including fluidization chamber of mild steel, 150 mm internal dia. and 250 mm height with the distributor plate having 1.5 mm holes and 35% free area, air chamber, Heater (2 KW rating), orifice meter, control valve and blower. Two fluidization chambers were fabricated for facilitating continuous drying operation at a later stage.
The proper design of the gas distributor plate makes the even air velocity across the fluidized bed directly relating to the bubble characteristics determining accordingly the proper mass and heat transfer as reported elsewhere (8).



1. Fluidization chamber 2. Air chamber 3. Heater 4. Orifice meter 5. Control valve 6. Blower

Figure2. Schematic diagram of Fluidized Bed Dryer for present work

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2. MODEL FOR BATCH FLUIDISED BED DRYER AND SIMULATION

In order to predict the drying characteristics based on the experimental data found from the above fabricated lab-scale set-up, the deterministic, stochastic or artificial intelligence may be resorted to be developed. The model has described the bed divided by solid phase, interstitial gas phase and bubble phase with the following assumptions:

1. Particles are homogenous, spherical and uniform.
2. In the bubble gas phase both the gas and bubble move upward whereas in the emulsion phase the solid particles move downward in the bed.
3. All particles are at the same temperature and having the same moisture content.
4. Drying medium in the fluidized bed is in thermal equilibrium with the particles.
5. Bubble size is uniform throughout the bed and drying in the freeboard is negligible.
6. Intra-particle moisture movement is described by the Fick's law.

The Fick's diffusion equation used for modeling the drying kinetics of the chickpeas in fluidized bed as dealt with by the present paper indicates the dependence of the evaluated effective diffusivity coefficient on solid hold up. The diffusion coefficient which is the important factor in the fluidized bed drying depends on temperature, concentration and the solid hold up and is estimated by Fick's diffusion equation. The lower drying constant value of chickpea attributed by its impervious skin has not been taken into account for the present study. The thermo-physical properties including moisture content would be of importance for heat and mass transfer in fluidized bed drying of chickpea.

Critical moisture content

Moisture diffusion rate from core of the particle to its surface equals to the rate of water evaporation from surface till critical moisture, i.e., constant rate drying ceases and falling rate starts which means the end of the constant rate period and the start of the falling rate period. It decreases with a decrease in initial moisture content of solid particle and also with an increase with solid hold up. The drying rate is influenced by the superficial velocity of drying medium. The chickpea solid particles are assumed to be uniform in spherical size and isotropic in nature.

Constant rate period

Moisture movement within the solid is rapid enough to maintain a saturated condition at the surface and the drying rate is controlled by the rate of heat transfer to the evaporating surface. The drying rate in the constant rate period is supposed to be controlled by the gas-particle heat transfer rate. External transport control is encountered in removal of free and weakly bound moisture and influencing factors for this model are temperature and velocity of heating medium, solid holdup. It is assumed that the gas leaving the fluidized bed need not necessarily be saturated and that the resistance for mass transfer lies around the film surrounding the particle. The drying rate in the constant rate period is increased by the temperature increase, although this effect is smaller in diffusion-controlled process. The following shell balance for moisture and enthalpy are written for the solids, interstitial gas phase, and the bubble phase. The moisture balance for the solids in the fluidized bed gives

$$\rho_s\left(1-\varepsilon_f\right)\frac{dC}{dt} = -W, \tag{1}$$

where, ρ_s is chickpea solid density, ε_f is void fraction in a bubbling bed as a whole and W is weight of chickpea in gm.

The bubble characteristics including their interchange coefficient on the basis of their axial position through the fluidized bed plays significant role contributing towards transport phenomena of the bed.

As δ is the bubble void fraction, ε_b is the volume fraction of the bubbles in the bed and ε_e is void fraction in the emulsion phase of a bubbling bed, the average bed voidage is given by

$$\varepsilon_f = \varepsilon_b\delta + (1-\varepsilon_b)\varepsilon_e \tag{2}$$

It has been reported that bubbles considered to be in plug flow with no axial dispersion may contain a small amount of solids, with negligible error; the bubble void fraction can be taken as unity. The present model assumes that the clouds surrounding the rising bubbles are such a thin in measurement that bubble phase exchanges heat and mass only with the interstitial gas phase. The emulsion phase may be assumed to be at minimum fluidization condition.

$$Hence,\quad \left(1-\varepsilon_f\right) = (1-\varepsilon_b)(1\text{-}\varepsilon_{mf}) \tag{3}$$

Substitution of this equation in equation (1) gives the moisture balance for chickpea in the dense phase leading to the following equation.

$$\rho_s\left(1-\varepsilon_{mf}\right)(1-\varepsilon_b)\frac{dC}{dt} = -W \tag{4}$$

The moisture balance for interstitial gases in the dense phase is given by

$$\rho_g\varepsilon_{mf}(1-\varepsilon_b)\frac{dY}{dt} + \rho_g m_d(Y_d - Y_i) + \frac{6K_{b\rho_g\varepsilon_b}}{d_b}(Y_d - Yb = W \tag{5}$$

where, ρ_g is fluid density,

The second and the third terms on the left hand side of the above equation (5) represent the net moisture gain by the gas and net heat transfer from the dense phase to the bubble phase respectively.

Combination of above equations (4) and (5) gives

$$\rho_g\varepsilon_{mf}(1-\varepsilon_b)\frac{dYg}{dt} + \rho_g m_d(Y_d - Y_i) = -\rho_s\big(1- \varepsilon mf1{-}\varepsilon bdCdt{+}6Kb\rho b\varepsilon bdbYb{-}Yd \tag{6}$$

The enthalpy balance for chickpea solid in the dense phase is given by

$$\rho_s\left(1-\varepsilon_{mf}\right)(1-\varepsilon_b)(\alpha_s + \alpha_w C)\frac{dT_s}{dt} = Q - \lambda W \tag{7}$$

where, λ is growth per cm., C denotes mean concentration, α term denotes thermal diffusivity, T_s solid chickpea temperature.

In the above equation ρs(1- ε_b) represent the solids fraction in the bed. The term on the left hand side of the equation (7) represents the rise in the sensible heat of the chickpea particle. Q represent the heat input, while W is the heat loss through evaporation of moisture from chickpea solid particle. The enthalpy balance for interstitial gas in the dense phase is given by

$$\rho_s \varepsilon_{mf}(1-\varepsilon_b)(\alpha_g + Y\alpha_y)\frac{dT_g}{dt} = \rho_g m_d(\alpha_g + Y_i\alpha_y)(T_i - Td+6hbb\varepsilon\varepsilon bdbTb-Td+Q \quad (8)$$

The first and the second term on the right hand side of equation (8) represent the enthalpy transfer, from the drying medium to the dense phase and from bubble phase to dense phase respectively. Combining equation 7 and 8 and substituting for W from equation (5) gives

$$\rho_s(1-\varepsilon_{mf})(1-\varepsilon_b)(\alpha_s + \alpha_w C)\frac{dT_s}{dt} = \rho_g m_d(\alpha_g + Yi\alpha yTi-Td-\rho gmd\alpha g+Yi\alpha yTi-Tdf-\rho g\varepsilon mf1-\varepsilon b\alpha g+Yi\alpha ydTgdt+6hb\varepsilon bdbTb-Td-\lambda\rho g\varepsilon mf1-\varepsilon bdYgdt+\rho gmdYb-Yi-6Kb\rho b\varepsilon bdbYb-Yd, \quad (9)$$

where, T_i is temperature of interstitial phase, T_d is dense phase temperature, K_b mass transfer coefficient in bubble phase; h_b is heat transfer coefficient in bubble phase, m is fraction of bed solids.

If the term involving the rate of change of the humidity and the temperature of the gas with time in the dense phase within the bed are assumed small compared to the corresponding convective terms, equations 5, 6 and 9 reduce to

$$\rho_g m_d(Y_d - Y_i) + \frac{6K_b\rho_g\varepsilon_b}{d_b}(Y_d - Y_b) = W \quad (10)$$

$$\rho_g m_d(Y_d - Y_i) = \rho_s(1-\varepsilon_{mf})(1-\varepsilon_b)\frac{dC}{dt} + \frac{6K_b\rho_g\varepsilon_b}{db}(Y_b - Y_d) \quad (11)$$

$$\rho_s(1-\varepsilon_{mf})(1-\varepsilon_b)(\alpha_s + \alpha_w C)\frac{dT_s}{dt} = \rho_g m_d(\alpha_g + Yi\alpha yTi-Td+6hb\varepsilon bdbTb-Td-\lambda[\rho gmdYd-Yi-6Kb\rho g\varepsilon b dbYb-Yd] \quad (12)$$

The moisture balance for the bubble phase is expressed as similar to equation (5) as

$$\rho_g\varepsilon_b\frac{dY_b}{dt} + \rho_g m_b(Y_b - Y_i) = \frac{6K_b\rho_b\varepsilon_b}{d_b}(Y_d - Y_b) \quad (13)$$

Neglecting the rate of change of humidity of the gas with time as compared to the corresponding convective terms, equation (13) becomes

$$\rho_g m_d(Y_b - Y_i) = -\frac{6K_b\rho_b\varepsilon_b}{d_b}(Y_d - Y_b) \quad (14)$$

By simplification,

$$Y_b = (6K_b\varepsilon_b Y_d + d_b m_b Y_i)/(6K_b\varepsilon_b + d_b m_b) \quad (15)$$

The enthalpy balance for the bubble phase can be expressed as similar to equation (8) as

$$\rho_g\varepsilon_b(\alpha_g + \alpha_y Y_i)\frac{dT_b}{dt} + \rho_s m_b(\alpha_g + Y_i\alpha_y)(T_i - T_b) = \frac{6h_b\varepsilon_b}{d_b}(T_b - T_d) \quad (16)$$

Neglecting the rate of change of the temperature of the gas with time when compared to the corresponding convective terms, equation (16) becomes

$$\rho_s m_b(\alpha_g + Y_i\alpha_y)(T_i - T_b) = \frac{6h_b\varepsilon_b}{d_b}(T_b - T_d) \quad (17)$$

from which, $T_b = \rho_g m_b(\alpha_g + Y_i\alpha_y)T_i d_b + 6h_b\varepsilon_b T_d/(\rho_g m_b(\alpha_g + Y_i\alpha_y)d_b + 6h_b\varepsilon_b)$ (18)

For a homogenous bed of (d=0)equations 11 and 12 reduce to

$$\rho_g \mathrm{m}_t(Y_0 - Y_i) = \rho_s(1-\varepsilon_f)\frac{dC}{dt} \quad (19)$$

$$\rho_s(1-\varepsilon_f)(\alpha_s + \alpha_w C)\frac{dT_s}{dt} = \rho_g m_t(\alpha_g + Y_i\alpha_y)(T_i - T0-\lambda[\rho gmtY0-Yi] \quad (20)$$

However, the above equations (19) and (20) refer to a special case of absence of the bubbles in the gas solid fluidized bed.

Equations(11) and(12) can be solved to obtain the humidity Y_b and Y_d and the temperature T_b and T_d of the gas leaving the fluidized bed, assuming that the gas humidity and the bed temperature correspond to saturation humidity Y_{sat} and wet bulb temperature referred to in the inlet air conditions. Since resistance to mass transfer lies in the film surrounding the solid chickpea particle, substitution of equation (21) in equation (10) gives the following expression

$$W = (1-\varepsilon_b)\rho_g a\, K_y(Y_{sat} - Y_d) \quad (21)$$

$$\rho_g m_d(Y_d - Y_i) - \frac{6K_b\rho_b\varepsilon_b}{d_b}(Y_b - Y_d) = (1-\varepsilon_b)\rho_g K_y\, a(Y_{sat} - Y_d) \quad (22)$$

The drying rate during the constant rate period is obtained by solving simultaneous equations (11), (12) and (22). Correct choices of the dense phase gas temperature, T_d and the dense phase gas humidity Y_d satisfy equation (12) with the left hand side set to zero with the inference that the temperature of the bed remains constant during the constant rate period. Using Y the drying rate $\frac{dC}{dt}$ is computed from equation (11).

A number of correlations are available in literature for predicting the bubble diameter, d_b, the mass transfer coefficient from the particle surface to the interstitial gas, K_y, u superficial velocity, D_e effective diffusivity, h is height of the column, the minimum fluidization velocity u_{mf} and the bubble rise velocity u_b required for use in equation 11, 12 and 20.In present model concerned the following Semiempirical correlations are used.

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For bubble size,

$$d_b = 0.474 \left[\frac{u-u_{mf}}{1.6{D_e}^{0.4} g^{0.5}}\right]^{0.4} (h + 3.94^{0.5})^{0.8} \quad (23)$$

For mass transfer coefficient from the dense phase to the bubble phase

$$K_b = \left[\frac{u_{mf}}{3}\right] + \left[\frac{4D_e \varepsilon_{mf} u_b}{\pi d_b}\right]^{0.5} \quad (24)$$

$$Ar = 3.142 * Dc^2/4 \quad (25)$$

Where Ar is the cross sectional area of the column,

v is the volumetric flow rateand Dc is the column diameter

$$u = v/Ar \quad (26)$$

Bubble rise velocity

$$u_b = 1.6{D_e}^{0.4}(g d_b)^{0.5} + (u - u_{mf}) \quad (27)$$

The minimum fluidization velocities

$$u_{mf} = \left[\frac{\mu_g}{\rho_g d_b}\right] [((33.7)^2 + 0.0408 \frac{{d_p}^3 \rho_g (\rho_s - \rho_g) g}{{\mu_g}^2})^{0.5} - 33.7] \quad (28)$$

The mass transfer coefficient from particle surface to the bulk gas

$$K_y = \left[\frac{D_e}{\Psi \Phi d_p}\right] [2 + 1.8{R_e}^{0.5} Sc^{0.33}] \quad (29)$$

Where R_e is Reynolds number at superficial velocity, S_c is Schmidt number.

For the heat transfer coefficient from the bubble phase to the dense phase

$$h_b = \left[\frac{u_{mf} \rho_g \alpha_g}{3}\right] + \left[\frac{4\rho_g \alpha_g K_g \varepsilon_{mf} u_b}{\pi d_b}\right]^{0.5} \quad (30)$$

Falling rate period

For tightly bound moisture within the particle, diffusion of moisture to the surface is limiting factor for drying operation. Rate of moisture removal drops with time and the internal movement of moisture, diffusion resistance inside the particle outweigh the film resistance. The drying rate is expected to be proportional to the free moisture content in the particles during the falling rate period. The estimated diffusion coefficient is usually a single kinetic parameter for assessing the drying kinetics and using the Fick's diffusion this is supposed to vary with other variable like solid hold up. This is based on capillary flow of liquid movement due to surface to surface force on the classical diffusion theory related to Fick's law.

The present paper assumes that moisture diffuses in the liquid phase to the chickpea solid particle surface across which evaporation occurs.

The moisture concentration is given by

$$\frac{\partial C}{\partial t} = D_e \left[\frac{\partial^2 C}{\partial r^2} + \frac{2\partial C}{r \partial r}\right] \quad (31)$$

The boundary conditions are:

At $t=0$; $0<r<R$, $C_i = C_o = C$

At $t>0$, $r=R$,if the sphere is initially at a uniform concentration C_1 and there is the surface condition

$$D_e \frac{\partial C}{\partial r} = K_y (C - C_0) \quad (32)$$

Where C is the actual concentration just within the sphere, and C_0 is the concentration required to maintain equilibrium with the surrounding atmosphere.

The possible solution by the Crank (4) equation for dimensionless moisture ratio

$$\frac{C - C_0}{C_1 - C_0} = \frac{2RN_{sh}}{r} \sum_{n=1}^{\infty} \frac{\exp\left(-D_e * \beta_n * \beta_n * \frac{t}{R^2}\right) * \sin(\beta_n * \frac{r}{R})}{\{{\beta_n}^2 + N_{sh} * (N_{sh} - 1)\} * \sin(\beta_n)} \quad (33)$$

where, Sherwood number $N_{sh} = \frac{dp\, K_y}{D_e}$

dp is chickpea particle diameter

β_n is the root of the equation

$$\beta_n \cot(\beta_n) + N_{sh} - 1 = 0$$

3. RESULT AND DISCUSSION

The above established mathematical equations describing the fluidized bed dryer have been resorted to using MATLAB programming. The following 6 figures (fig. nos. 3-8) depicting the trend of variation of $(C\text{-}C_0)/(C_1\text{-}C_0)$, i.e., dimensionless moisture ratio with different parameters like chickpea particle diameter, effective diffusivity, volumetric flow rate, fluidized bed column diameter give the deciding factors to the designer of the industrial fluidized bed dryer for optimum and economic storage and packaging of chickpea. The temperature dependence of diffusivity coefficient of chickpea with its activation energy value of 1238 KJ/Kg can be described by the Arrhenius-type relationship. Studying the following figures it is found that for all the chickpea particle diameters, by increasing the column diameter, dimensionless moisture ratio decreases. At a particular set of parameters including column diameter, particle diameter as the effective diffusivity decreases, the moisture ratio decreases. The trend of variation of the moisture ratio with the parameters has been shown in these following figures. The drying rate has been found to increase with increase of air fluidizing air flow rate and air temperature directly affecting the diffusivity. The results indicate that at low drying temperature at higher air velocity the drying performance would be more as indicated by the variation of the dimensionless moisture ratio.

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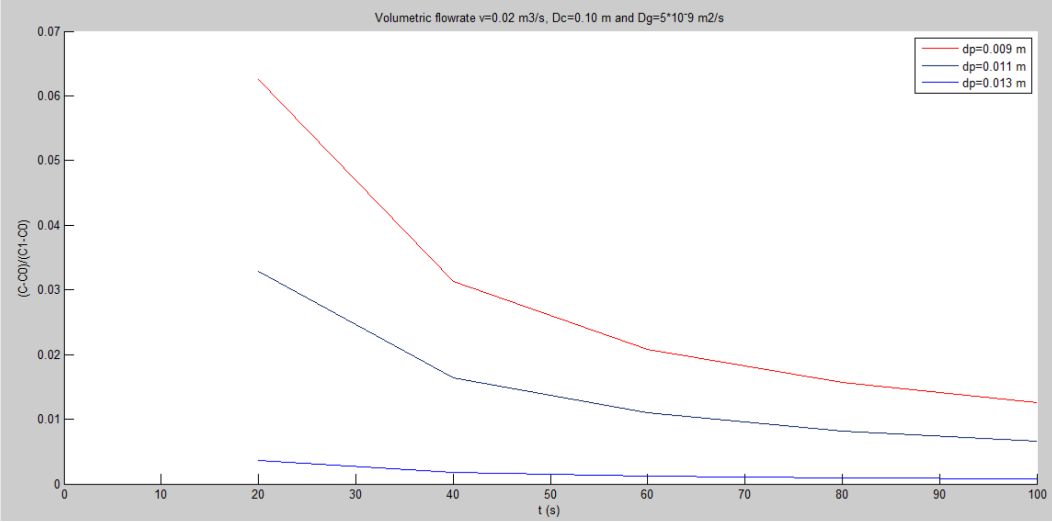

Figure 3: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.10 m, effective diffusivity 5×10^{-9} m^2/s and volumetric flow rate v=0.02 m^3/s.

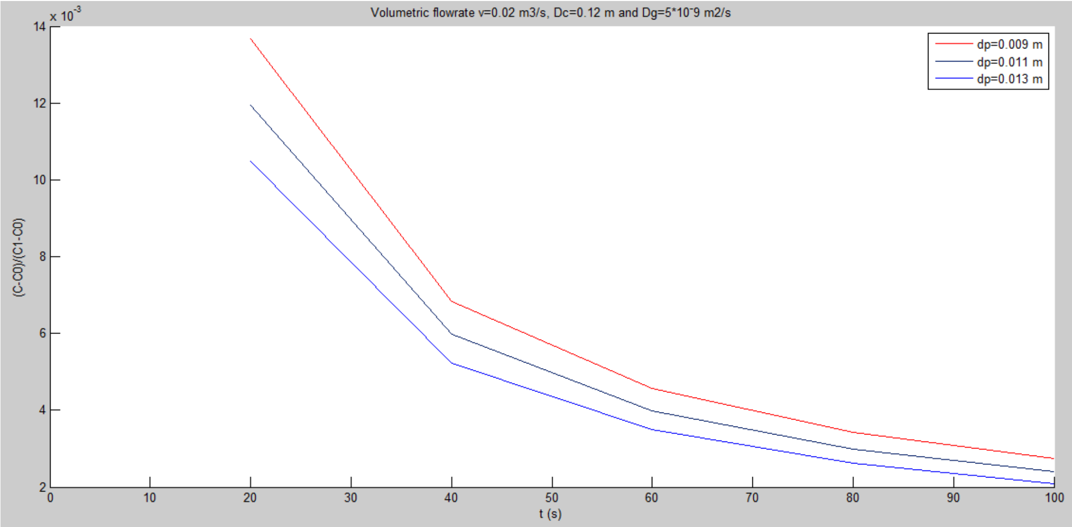

Figure 4: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.12 m, effective diffusivity $5\times10^{-9}m^2/s$ and volumetric flow rate v=0.02 m^3/s.

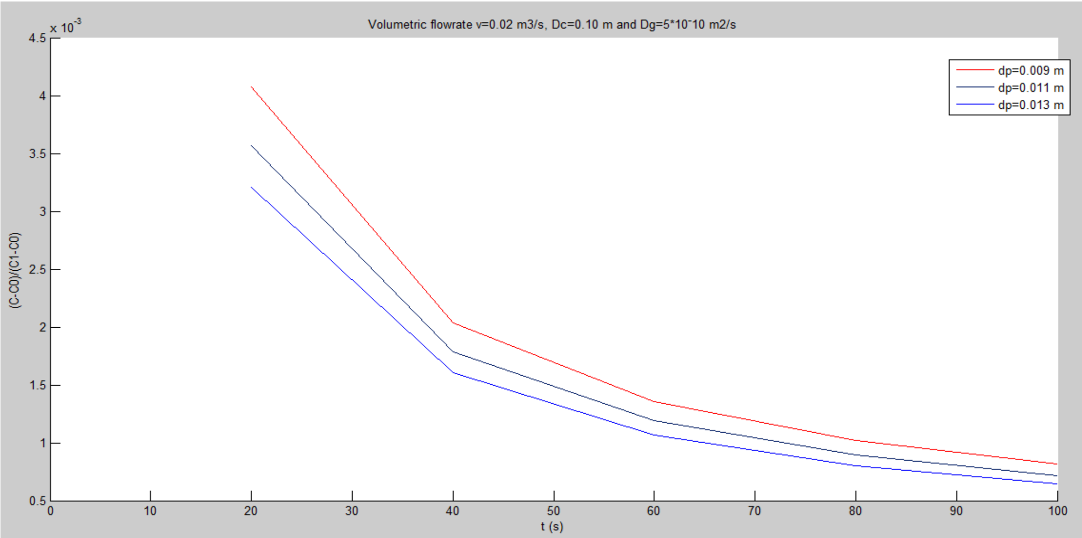

Figure 5: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.10 m, effective diffusivity $5\times10^{-10}m^2/s$ and volumetric flow rate v=0.02 m^3/s and 10 m^3/s.

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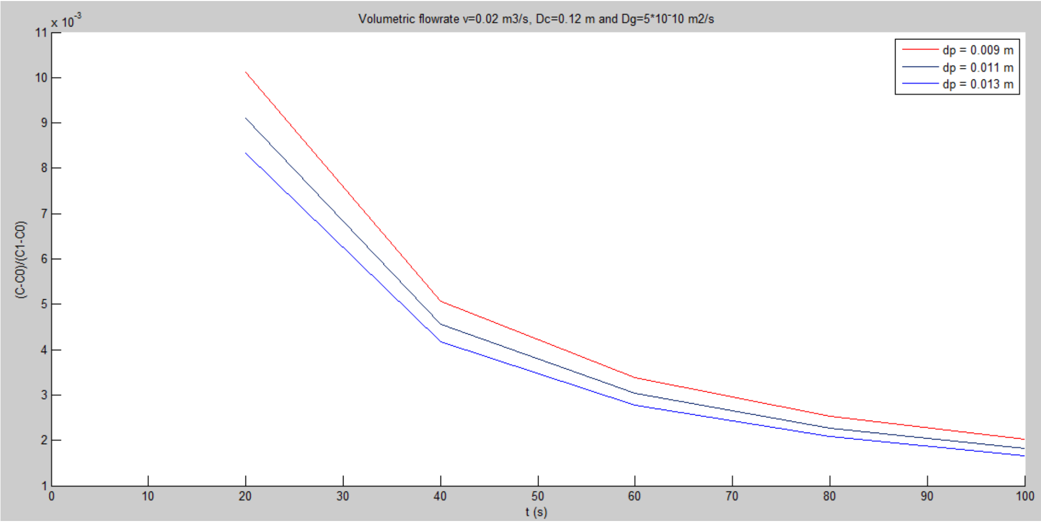

Figure 6: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.12 m, effective diffusivity $5\times10^{-10}m^2/s$ and volumetric flow rate v=0.02 m^3/s and 10 m^3/s.

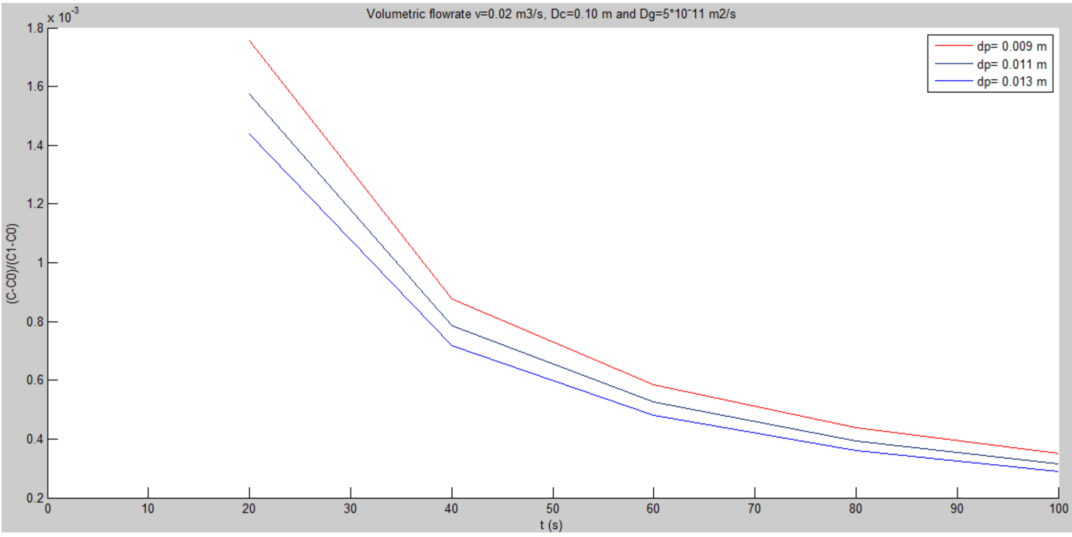

Figure 7: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.10 m, effective diffusivity $5\times10^{-11}m^2/s$ and volumetric flow rate v=0.02 m^3/s and 10 m^3/s.

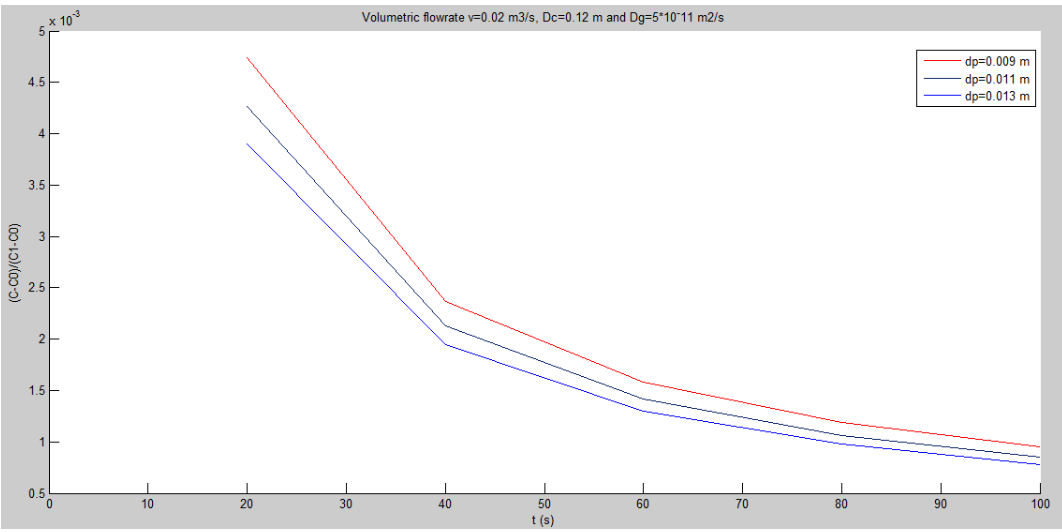

Figure 8: Drying for particle diameter dp=0.009, 0.011, 0.013 m, column diameter Dc=0.12 m, effective diffusivity $5\times10^{-11}m^2/s$ and volumetric flow rate v=0.02 m^3/s and 10 m^3/s.

4. CONCLUSION

The mathematical model has been made validated by comparing the simulated results using MATLAB programming with the experimental data including those found from literature. The inlet air velocity, in other form, volumetric flow rate was found to contribute much significant influence the effect on the dimensionless moisture ratio which gives the direct effect on the fluidized bed drying. A satisfactory agreement between the simulated data and the experimental data was achieved. From simulated data it was confirmed to ensure that the increase in solid hold up without getting the quality of fluidization affected. The selection and design of a particular fluidizing system would depend on the property of the material to be dried and the cost factor.

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