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Finite Volume Method for Solving Three-Phase Equations in Fluidized Bed

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--ABSTRACT

Three-Phase model making equations in fluidized bed dryers are mass and energy balance equations between these three phases. The purpose of this study is to solve the energy balance equations with finite volume method.
Governing equations are derived from Rizzi et al. model equations and have been solved simultaneously by using finite volume method. For descretizing equations upwind differencing scheme in space and explicit scheme for temporal differencing have been used. Finally a correlation for convection heat transfer coefficient between solids and interstitial gas phases has been presented. The Computational domain is chosen so that the numerical results can be compared with Rizzi et al. experimental data. The results of this study showed : there is a very good agreement between the numerical and experimental results and rate of heat transfer at the beginning of the fluidization is also very high.
KEY WORDS: Fluidized bed, Drying, Heat transfer coefficient, Three-Phase model, Geldart D.

1-INTRODUCTION

The Fluidized Bed Dryer is most suitable for drying granular, crystalline, pharmaceuticals materials, fine chemicals, dyes and food. It offers important advantages over other methods of drying. Some of them are: uniform temperature distribution throughout the product and high rate of heat and mass transfer.

Many models have been presented to describe the drying behavior in fluidized beds, including Two-Phase and Three-Phase models. Toomey and Johnstone (1952) [10] presented Two-Phase modeling, stating that 'all gas in excess of that necessary to just fluidize the bed passes through in the form of bubbles'. Kunni and Levenspiel [2] used this theory to model different operations for particles belonging to A and B groups of the Geldart classification. But further researches showed the Two-Phase modeling wasn't completely accurate. Greonewold and Tsotsas [3] (1997) used a three phase model including solid phase, bubble phase and suspension phase. In their model the gas fraction that formed bubbles was reduced. Palancz (1982) [6] presented a mathematical model for continuous drying process in a fluidized bed dryer. The model was a three phase model including solid, bubble and interstitial gas. Based on his model, the bubble phase was in a plug flow while interstitial gas and solid particles were perfectly mixed. Wildhagen et al. (2002) [9] considered a three phase model to describe the drying of porous alumina in a fluidized bed. The solid phase was perfectly mixed and interstitial gas and bubble phase were in plug flow. The energy loss through bed walls was neglected. Vitor et al. (2004) [8] used the same three phase model for drying process of Biomass tapioca belonging to Geldart group B. But in their study the energy loss was considered. The governing equations were energy and mass balances in three phases. Rizzi et al. (2009) [7] modified Vitor et al. (2004) model and applied it to describe the heat transfer in a fluidized bed of grass seeds, belonging to group D of Geldart classification. The assumption of no water evaporation simplified the governing equations, restricted them to energy balance equations. In their research a correlation to calculate heat transfer coefficient and a quantity of heat loss through bed walls were presented.

The aim of this work is to study the heat transfer mechanism in a fluidized bed dryer based on Rizzi et al. (2009) model. Vitor et al. (2004) and Rizzi et al. (2009) used the parameter estimation procedure, PSO and ESTIMA optimization codes to obtain the heat transfer coefficient between solids and interstitial gas phases along with DASSL computational code to solve the model equations. But in this work the equations have been solved by finite volume method and the correlation for heat transfer coefficient between solids and interstitial gas phases has been obtained by least square method. The results of numerical simulation are compared with Rizzi et al. (2009) experimental data.

The remainder of the paper is organized as follows : In Section 2 we introduce governing equations and describe experimental procedure and method of solution. Section 3 discusses about results obtained out of numerical modeling and the comparison with experimental data. Finally we conclude the paper in Section 4.

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2-MATERIALS AND METHODS

2-1Governing Equations

The interactions between three phases including solid, interstitial gas and bubble are shown in figure 1. As it can be seen there is no interaction between solid and bubble phase. Thus, heat and mass don't transfer between these phases.

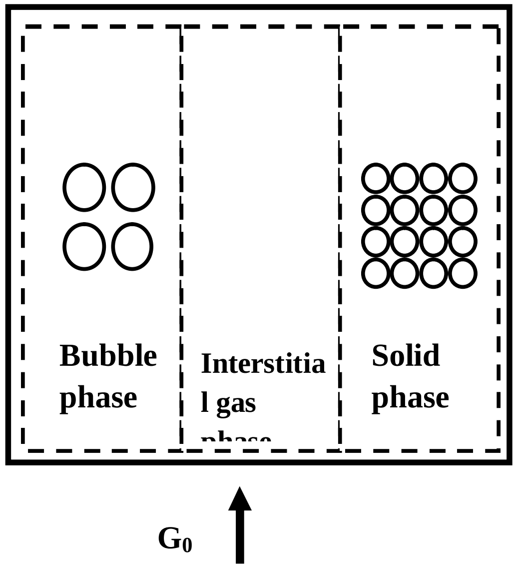

Figure 1: interaction between phases

The governing equations are simplified version of Vitor et al. model under the condition of no water evaporation. So governing equations are energy balances in three phases based on Rizzi et al. equations model.

In this model it is assumed the solid phase is perfectly mixed. Bubble phase moves as plug flow with no axial dispersion and interstitial gas phase can flow in an arbitrary flow regimes between the plug flow and the perfect mixing. In addition, heat transfer between interstitial gas and the dryer wall can occur, which means that this gas phase is the one responsible for heat loss.

Table (1) and Table (2) present model equations and Initial and boundary conditions respectively. The interactions between phases and heat loss are shown in Table (3). β_T is the coefficient that describes the gas regime along the bed height. Its value is given in Table (4). Table (5) represents the equations to calculate the properties of bed.

Table (1): Energy balance equations from Rizzi et al. model (2009) [7]

Phases	Equations	
Solid	$(1-\varepsilon)\rho_s \frac{dH_s}{dt} = f_{E1}$ $H_s = \left(c_{ps} - Y_s c_{pw}\right)(T_s - T_r)$	(1)
Interstitial gas	$(1-\delta)\varepsilon_{mf}\rho_g \frac{d\overline{H_\iota}}{dt} + G_{gi}\beta_T \frac{\overline{H_\iota}-H_0}{L} = f_{E2} - f_{E1} - E_W$ $\overline{H_\iota}=Y_g\lambda + \left(c_{pgi} + Y_g c_{pvi}\right)(\overline{T_{g\iota}} - T_r)$	(2)
bubble	$\delta\rho_g \frac{\partial H_b}{\partial t} + G_{gb}\frac{\partial H_b}{\partial z} = -f_{E2}$ $H_b = Y_g\lambda + \left(c_{pgb} + Y_g c_{pvb}\right)(T_{gb} - T_r)$	(3)

Table (2): Initial and boundary conditions [7]

Phases	I.C. and B.C.
Solid	$T_s(0) = T_{s0}$ $Y_s = constant = Y_s^*$
Interstitial gas and Bubble	$T_{gi}(0,z) = T_{gb}(0,z) = T_{s0}$ $T_{gi}(t,0) = T_{gb}(t,0) = T_{g0}$ $Y_g = constant = Y_g^*$

Table (3): Interaction between phases and heat loss [7]

	Equations	
Heat transfer between bubble and interstitial gas phases	$f_{E2} = h_b a_1 \left(T_{gb} - T_{gi}\right)$	(4)
Heat transfer between solid and interstitial gas phases	$f_{E1} = ha\left(\overline{T_{gi}} - T_s\right)$	(5)
Rate of heat loss through the column wall	$E_W = \alpha_w \frac{A_L}{V_{bed}}\left(\overline{T_{bed}} - T_{amb}\right) = \alpha_{wa}\frac{A_L}{V_{bed}}\left(\overline{T_w} - T_{amb}\right)$	(6)

Table (4): Values of β_T [7]

Flow regime	β_T
Perfect mixing	1.0
Plug flow	$\left(\frac{L}{\overline{H_i} - H_0}\right)\frac{\partial H_i}{\partial z}$
Arbitrary	$1.0 < \beta_T < 1.5$

Table (5): Equations to calculate properties of bed

	Relations	
Bubble diameter	$d_b = 2.25 z^{0.81} \times (u_0 - u_{mf})$	[7]
Bubble velocity	Werther relation: $u_b = \psi(u_0 - u_{mf}) + \alpha u_{br}$	[4]
Minimum fluidization velocity	$\left(\frac{1.75}{\varepsilon_{mf}{}^2\varphi_s}\right) Re_{mf}{}^2 + \left(\frac{150(1-\varepsilon_{mf})}{\varepsilon_{mf}{}^3\varphi_s{}^2}\right) Re_{mf} = Ar$	[5]
Minimum fluidization height	$L_{mf} = 4M_s/(1-\varepsilon_{mf})\rho_s \pi D_c{}^2$	[7]
Volumetric bubble concentration	$\delta = 1.0 - L_{mf}/L$	[7]

The physical properties of grass seeds used in Rizzi experiments are shown in Table (6):

Table (6): Physical properties of the seed

Name	Value	Unit
d_p	2.23×10^{-3}	mm
c_p	428	J/Kg K
ρ_p	1018	Kg/m^3
φ	0.92	--

2-2Experimental Procedure

The experimental setup is shown in Figure 2. Experiments are conducted in a glass cylinder of 0.07 m in diameter and 0.40 m in height and thickness of 0.005 m. Air is used as fluidizing gas and passes through a mesh filter, a 2 KW electrical heater and a mesh plate distributor, before fluidizing the bed of particles [7].

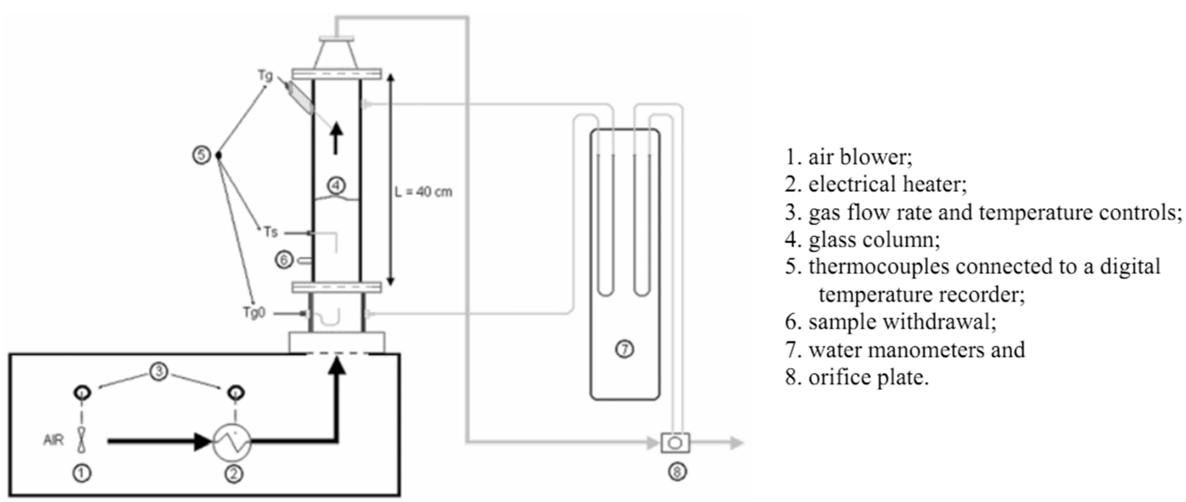

Figure 2: Experimental setup [7]

Three thermocouples are placed inside the column to measure: the inlet air temperature, T_{g0} ; the solid temperature, T_s ; and outlet air temperature, T_{gL} , (Thermocouple placed at height of 0.38 from the base of the column). The value of wall temperature is equal to the time-average from measurements obtained at height of 0.38.

Another thermocouple, located close to the experimental unit, registers the ambient temperature, Tamb , for each experiment. It's necessary to mention that the inlet gas temperature is assumed to be steady. Table (7) shows operational condition of experiments:

Table (7): Operational condition of experiments [7]

Test #	$G_g (kgm^{-2}s^{-1})$	T_{g0}(°C)	T_{s0}(°C)	T_{amb}(°C)	T_w(°C)	L(m)
1	0.948	51.4	26.2	27.3	33.9	0.182
2	1.301	30.9	18.1	19.6	22.3	0.227
3	1.075	40.9	26.4	29.2	32.8	0.204
4	1.088	40.9	21.6	24.4	29.4	0.212
5	1.259	51.5	22.6	25.1	33.5	0.235
6	1.073	40.7	25.4	27.3	30.8	0.200
7	1.001	30.8	17.1	17.1	20.7	0.189
8	1.198	51.1	27.2	29.2	37.2	0.220
9	1.064	41.3	28.1	29.9	32.2	0.205

2-3 Method of solution

The convection heat transfer coefficient between solid and interstitial gas phases is expressed below as a function of Reynolds number:

$$h = \frac{k_{gi}}{d_p}\left(x_1 {Re_p}^{x_2}\right) \qquad (7)$$

By using solid phase equation for each group of experimental data, a quantity has been estimated for heat transfer coefficient between solid and interstitial gas phases. Therefore by having values of h and Re_P for each experiment along with using the least square method, the unknown parameters (x_1 and x_2) of heat transfer equation will be calculated.

Finite volume method has been used to solve equations. First the flow field in the bed was divided into a finite number of control volumes after considering the bed expansion. The energy equations in three phases were descretized by using upwind differencing scheme in space and Explicit scheme for temporal differencing. Therefore the differential equations were converted to algebraic equations. Finally by solving these algebraic equations simultaneously, temperature distributions will be achieved.

3-RESULTS AND DISCUSSION

After calculating the unknown parameters of convection heat transfer coefficient (7), the equation has been obtained as below:

$$h = \frac{k_{gi}}{d_p}\left(2 \times 10^{-4} {Re_p}^{1.105}\right) \qquad (8)$$

As mentioned before, to validate the numerical results, Rizzi et al. experimental data has been used. These experimental data are the variation of solid and outlet gas temperatures with time. In Figures (3)-(5) comparison of experimental data and the numerical results are presented.

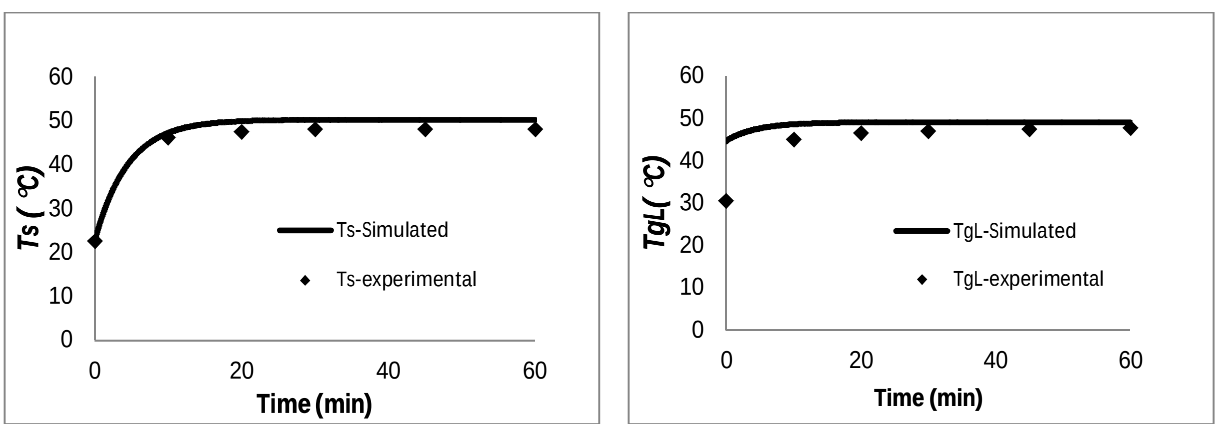

Figure 3: Comparison between experimental and simulated data for test#5

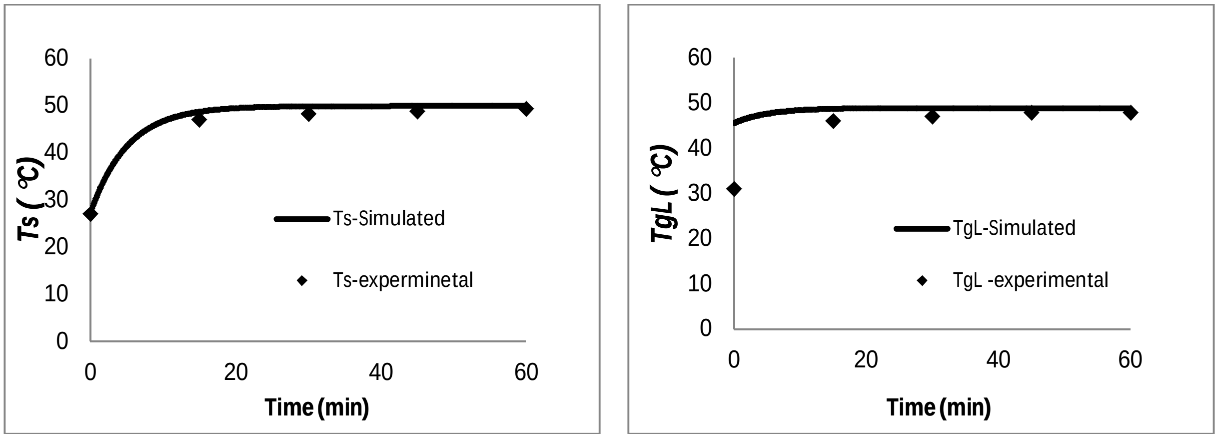

Figure 4: Comparison between experimental and simulated data for test#8

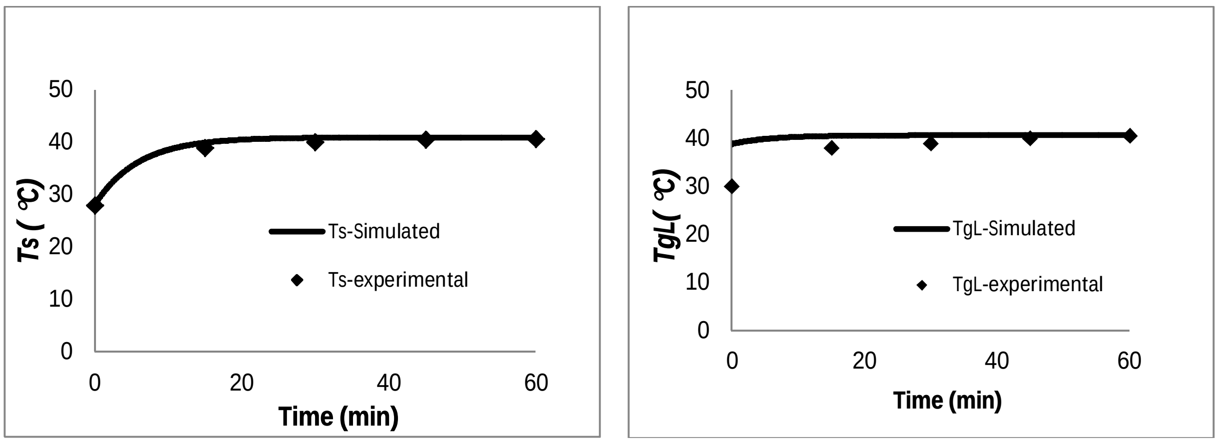

Figure 5: Comparison between experimental and simulated data for test#9.

It's seen from Figures 3-5 that maximum slope of temperature variations occur at the beginning of fluidization process which show high rate of heat transfer. It's true due to large temperature differences between phases in the beginning instants. Also the speed of solid particles temperature variation is more than variation related to outlet gas but after transient time (after t>15 minutes) they inclined to each other.

4-Conclusion

In this study, Rizzi et al. (2009) model has been used to describe heat transfer phenomena in fluidized bed of grass seeds belonging to Geldart group D. and a new correlation was proposed to calculate heat transfer between solid and gas phases. A suitable numerical method that employs finite volume method has been applied to discretize the equations. Results obtained out of numerical modeling were compared with experimental data in Figures (3)-(5). And as it can be seen in these Figures the experimental and numerical results are in general in good agreement.

5-Nomenclature

A_L	lateral area of the bed column = $\pi D_c L$	m^2
a	specific exchange superficial area between gas and particles	m^{-1}
C_p	specific heat	J/Kg.K
D_c	column diameter	m
E_w	rate of energy loss through column wall per unit of bed volume	J/s
G	superficial mass flow rate	$Kg/m^2.s$
h	heat transfer coefficient between solid and interstitial gas	$J/m^2.s.K$
H	specific enthalpy	m^2/s^2

$h_b a_1$	volumetric heat transfer coefficient between bubble and interstitial gas	$J/m^3.s.K$
L	expanded bed height	m
M	mass	kg
T	temperature	℃ or K
t	time	s
V	volume	m^3
Y	water content in dry basis	(-)
Y*	equilibrium moisture content	(-)
z	axial coordinate	m
Greek symbols		
α_w	heat transfer coefficient between column walls and air ambient	$J/s.m^2.K$
β	coefficient related with the interstitial gas phase	(-)
δ	volumetric bubble concentration	(-)
ε	bed porosity	(-)
φ	particle sphericity	(-)
λ	latent heat of vaporization	J/Kg
µ	viscosity	Kg/s.m
ρ	density	Kg/m^3
ψ	ratio of the visible bubble flow to the excess gas velocity	(-)
Subscripts		
0	initial value	
amb	ambient	
bed	bed	
b	bubble	
g	gas	
gL	exit gas	
I	interstitial gas	
mf	minimum fluidization	
p	particle	
s	solid	
v	water vapor	
w	wall	

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