Improving Three Phase Modeling of Fluidized Bed Dryer

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ABSTRACT: *Heat transfer phenomena in batch fluidized bed dryer have been studied and analyzed using three phase model including solid phase, interstitial gas phase and bubble phase based on Rizzi model. New correlations of heat transfer coefficient to surrounding and convection heat transfer coefficient between solid and interstitial gas are being proposed. In addition, some modifications have been done according to the state of flow and particle properties.The model was used to study the effects of particle diameter, particle density and input gas temperature on Geldart group D particles. The numerical results of heat transfer such as solid and gas temperature distributions have been compared with the available experimental data and good agreement between them is observed.*

KEY WORDS: *Three phase modeling, Drying, Fluidized bed, Heat transfer coefficient, Geldart D.*

INTRODUCTION

The fluidized bed dryer is used extensively in a wide variety of industries because of its large capacity, low construction cost, easy operability, and high thermal efficiency [1]. The most important advantage of these dryers is high heat and mass transfer coefficient between gas and particles due to good contact between them and also due to intensive solid mixing. According to these advantages, study and optimization of these dryers is so essential. Study of heat transfer dryers has a long history started by two phase modeling in 1952. The base assumption of two phase theory was that ‘all gases in excess of that necessary to fluidized the bed passes through the bed in the form of bubbles.’ *Kunni & Levenspiel* [1] (1996) used this theory to model different operations for particles belonging to Geldart group A and B. As *Yates* [2] (1983) stated, the serious challenge is the amount of gas that this theory assumed in the form of bubble in the bed. *Tsotsas & Groenewold* [3] (1997) by using a three phase modeling of fluidized bed dryers, including solid phase, bubble phase and suspension phase, decreased the volume fraction of bubble in their models. *Palancz* [4] (1982) presented a mathematical model for continuous drying process in a fluidized bed dryer, his model was a three phase model including solid, interstitial and bubble phase. He assumed the dilute bubble phase in a plug flow while the interstitial gas as well as the solid particles is considered to be perfectly mixed. *Wildhagen* [5] (2002) also described the drying process through a three phase model. The solid phase was considered as lumped and interstitial and solid phase

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Table 1: Energy balance equation of Rizzi et al. model (2009).

Phases	Energy balance equations	B.C. and I.C.
Solid	$(1-\varepsilon)\rho_s \frac{dH_s}{dt} = F_{E1}$ (1)	$T_s(0) = T_{s0}$ Y_s = constant = Y_s^*
Interstitial gas	$(1-\delta)\varepsilon_{mf}\rho_g \frac{d\bar{H}_1}{dt} + G_{gi}\beta_T \frac{\bar{H}_1 - H_0}{L} = f_{E2} - f_{E1} - E_W$ (2)	$T_{gi}(0,z) = T_{s0}$ $T_{gi}(t,0) = T_{g0}$ Y_g = constant = Y_s^*
Bubble	$\delta\rho_g \frac{\partial H_b}{\partial t} + G_{gb}\frac{\partial H_b}{\partial z} = -f_{E2}$ (3)	$T_{gb}(0,z) = T_{s0}$ $T_{gb}(t,0) = T_{g0}$ Y_g = constant = Y_s^*

was assumed as a perfect plug- flow. The governing equations were heat and mass transfer equations in three phase model. In this research, heat lost through the wall was ignored. *Vitor et al.* [6] (2004) investigated drying process of tapioca belongs to Geldart group D particles, in a batch fluidized bed by the purpose of defining heat and mass transfer coefficient between solid and gas. They did experimental and numerical modeling of the process. Vitor added heat lost through the wall to the heat transfer equations. *Rizzi et al.* [7] (2009) modeled drying process of grass seeds in a same manner as *Vitor et al.* the seeds belong to Geldart group D particles. In *Rizzi* research the assumption of no water evaporation has simplified the governing equations and the restricted them to energy balance equations. A parameter estimation method has been used in *Rizzi's* research and a quantity of heat lost coefficient and a correlation to calculate heat transfer coefficient has been presented.

In this study we want to add some modifications to *Rizzi et al.* model. One of them is using Nonaka and Horio relations to calculate bubble diameter that can use for a wide range of particle diameter and gas velocity. According to this choice our prediction of fluidization regime is more trustable, as the result quantity of ratio of the visible bubble flow to the excess gas velocity (ψ) can be choosing better. The other purpose of this work is to present a new correlation for convection heat transfer coefficient and heat lost coefficient through the wall. In new correlations characteristics of fluidized bed can represent better. The results of numerical simulations that concluded by using this new correlations compare to the result that concluded by using *Rizzi et al*. correlation. *Rizzi et al.* experimental data is used for validating of results. In this research least square method is used to calculate the coefficient of relations to estimate heat transfer coefficients.

THEORITICAL SECTION

Governing equations

Governing equations are the equations of heat balance for three phases including solid, interstitial and gas bubble phase. These equations are according to *Vitor et al.* equations to model drying process. The assumption of no water evaporation simplifies these equations. So the governing equations are Rizzi et al. model equations.

Assumption of model

The assumptions of the model are described as follows: (a) the energy losses on the dryer wall only occur in the interstitial gas phase; (b) the solid phase behaves as a perfect mixer; (c) all the transfer mechanisms presented in the bubble gas phase are purely convective and unidirectional; (d) there is no mass or energy transfer between solid and bubble gas phases

Model equations and initial and boundary conditions

Equations and initial and boundary conditions has been presented in Table 1 there is some terms in this table belong to interaction between phases; these terms have been described in Table 2. Beside energy balance equations, equations of enthalpy to temperature for each phase are needed to solve the equations and gain temperature distribution at any instant. β_T Coefficient is a parameter belongs to the assumption of gas flow in the bed. Different quantities of this coefficient have been shown in Table 3.

Table 2: Interaction between phases and heat loss to ambient.

Heat transfer between bubble and interstitial gas phase	Heat transfer between solid and interstitial gas phase
$f_{E2}=h_b a_I (T_{gb}-T_{gi})$ (4)	$F_{E1}=ha\left(\bar{T}_{gI}-T_s\right)$ (5)
Heat loss through the wall	
$E_W=\alpha_w \frac{A_L}{V_{bed}}\left(\bar{T}_{bed}-T_{amb}\right)=\alpha_{wa}\frac{A_L}{V_{bed}}\left(\bar{T}_w-T_{amb}\right)$ (6)	

Table 3: Quantity of β_t coefficient in energy balance equation.

Plug flow	Arbitrary flow	Perfect mixing
$\beta_t=\left(L/\left(\bar{H}_l-H_0\right)\right)\partial H_i/\partial z$	$1.0<\beta_t<1.5$	$\beta_t=1.0$

Table 4: Equations to calculate properties of bed.

Bubble diameter	Nonala et al. [9]
Bubble velocity	Werther $u_b=\psi(u_o-u_{mf})+\alpha u_{br}$ [5]
Minimum fluidization height	$L_{mf}=M_s/\left((1-\varepsilon_{mf})\rho_s A_b\right)$ [7]
Minimum fluidization velocity	$\left(1.75/\varepsilon_{mf}^3\phi_s\right)Re_{mf}^2+\left(150(1-\varepsilon_{mf})/\varepsilon_{mf}^3\phi_s^2\right)Re_{mf}=Ar$ [5]
porosity	$E=1-\frac{L_{mf}}{L}(1-\varepsilon_{mf})$ [7]
Volume fraction of bubble	$\delta=1-(L_{mf}/L)$ [5]

Method of solution

For each group of available experimental data a quantity has been estimated for convection heat transfer coefficient between solid and interstitial gas phase and a correlation like below represented for heat transfer coefficient.

$$h=\frac{k_{gi}}{d_p}\left(x_1 Re_p^{x_2} Ra^{x_3}\right) \tag{7}$$

After calculating coefficient of above equation by a least square method based on experimental data and solid phase equation, equations discretize in a finite difference manner and differential equations convert to a system of algebraic equations for three phases. After that, by using Rizzi experimental data for outlet gas temperature and reducing error of numerical results, quantity of heat loss coefficient estimated. By Calculating coefficient of below equation, new correlation of heat loss coefficient between interstitial gas and ambient has been achieved.

$$\alpha_w=x_1 Re_p^{x_2} Ra^{x_3}\alpha_d^{x_4} \tag{8}$$

At last this relation will be used in model and by solving the equations in a finite difference method, temperature distribution will be achieved. The effect of diameter and inlet gas temperature will be studied.

RESULTS AND DISCUSSION

Validations

This study has been done to model and investigate heat transfer phenomena in a fluidized bed of group D particle. A method to estimate heat transfer coefficients has been presented and new correlations describe properties of the bed better. One of advantages of this model is taking into account the variation of physical properties of gas to temperature. Respective equations have been shown in Table 4. Nonaka and Horio relations have been used to calculate bubble diameter. As it was said, the advantage of these relations is that, they can use in a wide range of particle diameter and gas velocity. In Fig. 1 the variation of bubble diameter for three set of experiment condition of Rizzi et al., that has been estimate by using Nonaka and Horio relations in the height of the bed has been drawn. According to Fig. 4 and

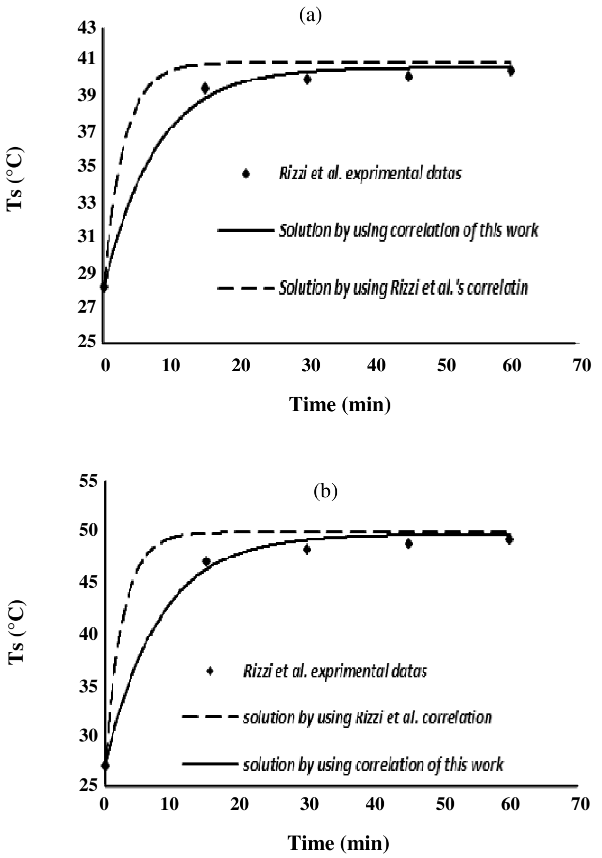

Fig 1: Comparison of two set of numerical results and Rizzi et al. experimental data. For solid temperature (a)test 8 of Rizzi et al. (b) test (9) of Rizzi et al.

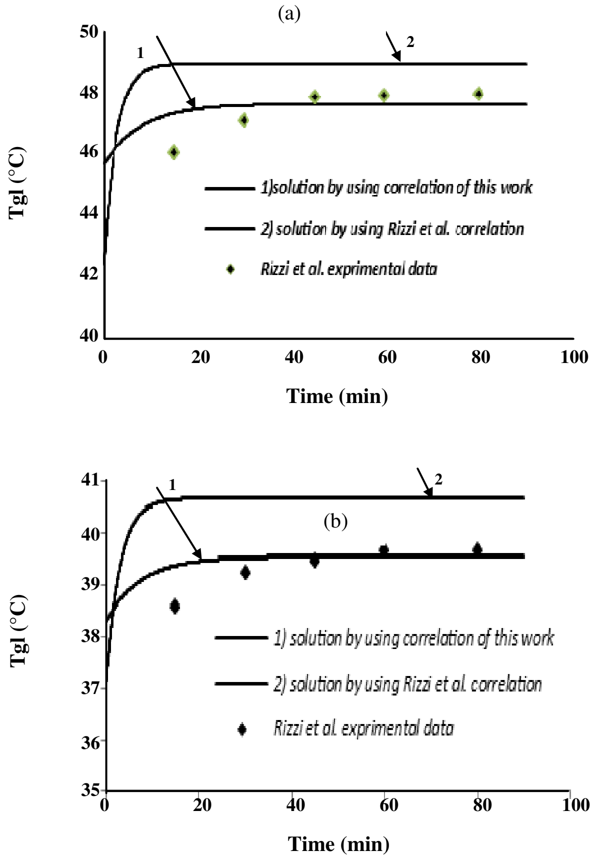

Fig 2: Comparison of two set of numerical results and Rizzi et al. experimental data for outlet gas temperature (a)test 8 of Rizzi et al. (b) test (9) of Rizzi et al.

the bed diameter (7cm), it can be concluded that a bubbling regime should supposed to exist in almost entire bed. This justifies the choice of ψ=0.26 for geldart group D particle in a bubbling regime according to Werther and Hiligaurd offer. Correlations to calculate heat transfer and heat loss coefficient has been obtained as below.

$$h = \frac{k_{gi}}{d_p}\left(0.7 \times Re_p^{0.03}\, Ra^{-0.2}\right) \qquad (9)$$

$$\alpha_w = 1.6 \times Re_p^{\frac{-9}{39}}\, Ra^{0.49} \alpha_d^{-2.67} \qquad (10)$$

To validate numerical results, *Rizzi et al.* experimental data has been used. Experimental data are distribution of temperature of solid and outlet gas temperature. The detail of these set of test has been explained in [8].

In Figs. 1 & 2 comparison of experimental and numerical results is observed. As it seen in Fig. 1 the result of distribution of solid temperature by using correlation of this work has better agreement to experimental data. The main difference between obtained chart by using Rizzi correlation and the other one concluded by using new correlation of present work is the slop of charts in beginning instants. So investigation of behavior of process in the beginning by experimental studies is very essential. Distribution of outlet gas temperature is shown in Fig. 2 and it observes that the correlation of this work in comparison with concluded results by using Rizzi correlation, has presented better results. Difference between slope of obtained charts in the beginning is also observed. More accurate experimental investigations specify if plug flow assumption need modification or not.

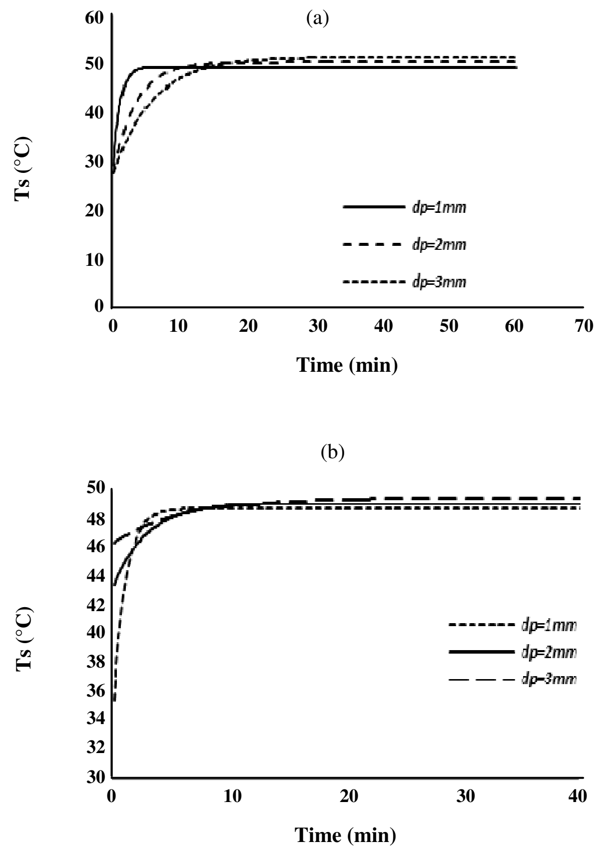

Fig 3: Effect of particle diameter on temperature distribution of (a) solid (b) outlet gas.

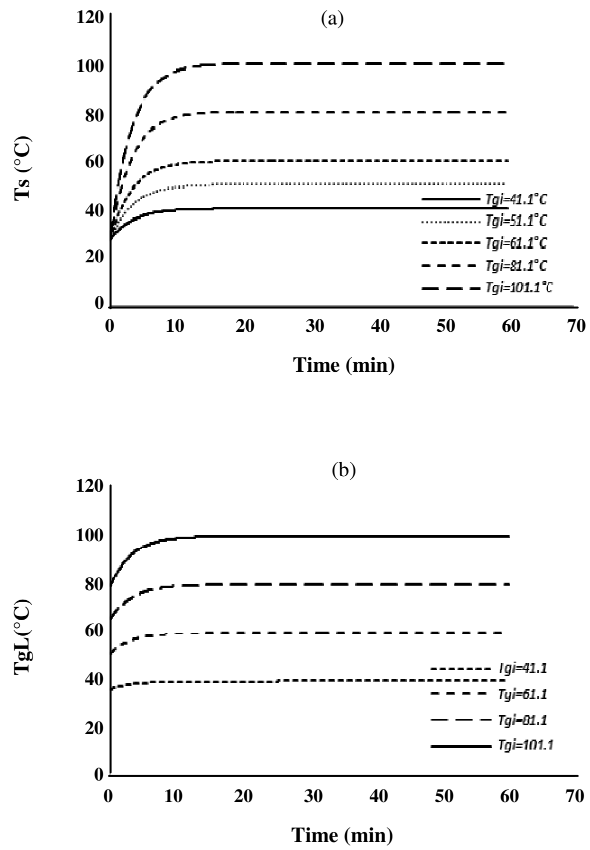

Fig 4: Effect of inlet gas temperature on temperature distribution of (a) solid (b) outlet gas.

Effect of variation of parameters

In Fig. 3 effect of particle diameter on temperature distribution of solid temperature to time has been shown. It is obvious in this figure that the slope of temperature line diminishes by increasing particle diameter and it means heat transfer decline. This behavior of particulate particles in fluid bed is justifiable by investigation of variation of physical properties of bed by particle diameter. Some of these variations are decreasing porosity and height of the bed in minimum fluidization mood. On the other hand volume fraction of bubble diminishes in the bed. All of this, causes a decrease in region of heat transfer, in other words, interaction between solid and gas diminishes and so temperature gradient has a less slope.

As the result of heat transfer rate diminishing, for larger particle the range of temperature variance for outlet gas temperature is lesser.

Effect of inlet gas temperature on heat transfer of particulate solid belongs to Geldart group D particle has been studied as shown in Fig. 4. It is observed that when inlet gas temperature increases, rate of heat transfer also have an increasing manner. It was predictable according to decrease of volume fraction of bubble and increase of mean bubble velocity for higher inlet gas temperature.

CONCLUSIONS

In three phase modeling that has been used in this paper some modifications have been added to *Rizzi et al.* model. Using *Horio & Nonaka* relations to calculate bubble diameter predicts the regime of fluidization better. And so visible volume fraction of bubble (ψ) has been chose better. Variation of physical properties of gas has taken into account and so the accuracy of the model increased. New correlations that have been presented to calculate convection heat transfer coefficient and heat loss coefficient of gas to ambient is more compatible to the base of convection heat transfer and characteristics of gas and bed wall. Comparison of numerical results

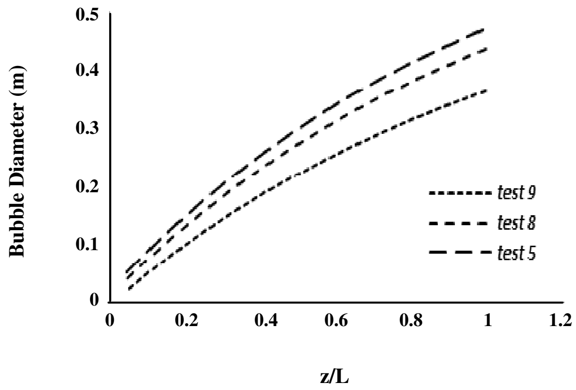

Fig 5: Variation of bubble diameter along height of the bed for three set of Rizzi et al. experimental data.

to available experimental data shows good agreement and it is observable that the results concluded by using new correlations of this work has more agreement to experimental data than results by using *Rizzi et al.* correlations. The more difference of this two obtained temperature line is at the beginning of the process, so it is essential to investigate the heat transfer process at these beginning instants. The other case that should be verified is the assumption of plug flow at the beginning of the process.

Effect of inlet gas temperature and particle diameter has been investigated and justifiable results show validity of the model and are good references to study the behavior of fluidized bed in heat transfer.

Nomenclature

a	Specific area of the interstitial gas-solid interface, m^{-1}
A_C	Cross sectional area, m^2
A_L	Lateral area of the bed column, m^2
Cp	Specific heat at constant press, J/kg.K
d_p	Particle diameter, m
G	Superficial mass flow rate, $kg/m^2.s$
h	Effective heat transfer coefficient between, $W/m^2.K$ Solid and interstitial gas phase
H	Specific enthalpy, J/kg
K	Thermal conductivity, W/m.K
L	Height of the bed, m
Ra	Rayleigh Number, Non-dimensional
Re_p	Reynolds number, Non-dimensional
T	Temperature, °C
T_{abs}	Temperature, K
M	Mass, kg
u	Velocity, m/S
u_{br}	Rise velocity of a bubble, m/s
Y	Moisture content in dry base, Non-dimensional
Y_s^*	Equilibrium moisture content in dry base, Non-dimensional
z	Axial coordinate in bed, m

Greek letters

β_T	Coefficient related with interstitial gas phase, Non-dimensional
δ	Bubble porosity, Non-dimensional
ε	Bed porosity, Non-dimensional
μ	Dynamic viscosity, $N.s/m^2$
ρ	Density, kg/m^3
ϕ	Sphericity, Non-dimensional
ψ	Ratio of the visible bubble flow to the excess gas velocity, Non-dimensional

Subscript

0	Initial conditions
amb	Ambient
b	Bubble, bed
s	Solid
v	Water vapor
w	Wall
g	Gas
gl	Exit gas
i	Interstitial
mf	Minimum fluidization
p	Particle

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