EXPERIMENTAL STUDY OF THE CONVECTIVE HEAT TRANSFER COEFFICIENT IN A PACKED BED AT LOW REYNOLDS NUMBERS

by

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An experimental study to evaluate the convective heat transfer coefficient in a cylindrical packed bed of spherical porous alumina particles is investigated. The task consists in proposing a semi-empirical model to avoid excessive instrumentation and time consumption. The measurement of the bed temperature associated to a simple energy balances led to calculate the gas to particle heat transfer coefficient using a logarithmic mean temperature difference method. These experiments were performed at atmospheric pressure. The operating fluid is humid air. The gas velocity and temperature ranged from 1.7-3 m/s and 120-158 °C, respectively. The data obtained was compared with the correlations reported in the literature. It is shown that the proposed model is in reasonable agreement with the correlation of Ranz and Marshall. Despite, many researches on experimental investigations of heat transfer coefficient in packed beds at low and average temperature are proposed, few studies presented calculation of convective heat transfer coefficient at high temperature (above 120 °C). A possible application of the proposed model is drying and combustion.

Key words: *heat transfer coefficient, energy balance, packed bed, temperature measurements*

Introduction

Packed beds have attracted a variety of applications, such as drying and chemical industries. Knowledge of heat transfer characteristics and temperature distributions in packed beds is of great importance. The particles to gas heat transfer coefficient is a key parameter to quantify heat transfer. In fact, a considerable amount of study has been carried out to evaluate this coefficient. Wakao *et al*. [1] provide a comprehensive review of the evaluation of the particle to fluid heat transfer coefficient. Ranz and Marshall [2] proposed the well-known correlation of this coefficient for spherical particle on the basis of a wide variety of experimental data on the evaporation of liquid droplets in air. Bird *et al*. [3] gave a set of heat transfer correlations for a

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packed bed of spherical and cylindrical particles. Rowe and Claxton [4] presented a Nusselt number correlation by measuring experimentally heat and mass transfer between the sphere and identical fixed spheres. Jingzhu *et al*. [5] measured particle to gas heat transfer coefficient from one-shot thermal responses in packed beds of glass beads air system in the range of Reynolds number for 5 to 230 and temperature were below 100 °C. Based on steady-state conditions, Galloway and Sage [6] developed a correlation for the gas to particle heat transfer coefficient for spherical pellets. Nasr *et al.* [7] gave an experimental study of forced convection heat transfer from a cylinder in a packed bed of spherical particles. Aluminum, alumina, glass, and nylon were used as packing materials. The operating fluid is air. The authors demonstrated that the uncertainly in the reported Nusselt numbers was ±20%. Balakrishnan and Pei [8] demonstrated that conduction between the particles in the bed and convection between the flowing gas and the particles which interact with each other, are the major reason for the difficulty in obtaining a single generalized experimental correlation or theoretical or semi-empirical models to evaluate the heat transfer coefficient in packed beds. Littman and Sliva [9] showed a strong dependence on Reynolds numbers especially for packed beds since the region near the point of contact between particles was not fully accessible to the flowing fluid.

The heat transfer coefficient was also determined for agro alimentary products: Wang *et al.* [10] and Bala [11] studied, experimentally, the convective heat transfer coefficient in packed bed of rice and malt, respectively. Khandker and Woods [12] measured the heat transfer coefficient in packed beds of barley. This coefficient was found to be a function of air flow rate.

Most of the previous studies presented above required excessive instrumentation and time.

Besides, the major correlations proposed in the literature are function of dimensionless numbers. Nevertheless, the heat transfer coefficient is shown to be affected by other parameters such as particles and bed characteristic and boundary conditions. Then, the application of a general correlation for packed beds is not possible. Moreover, the usual studies are at low and average temperatures. Hence, an attempt is made in this paper to quantify the heat transfer coefficient basing on temperature experimental data associated to simple energy balances at high temperature and low Reynolds numbers.

Experimental set-up

The experimental device consists of a thermally isolated cell containing the experimental packed bed, a steam generator, a hot gas heating system, a closed loop for the circulation of humid air (P_v = 0.6 bar), a data acquisition and control unit (fig. 1). The circulation of humid air is ensured by a fan. Gas flow rate and velocity in the cell are controlled by adjusting the speed of the fan. A control system allows setting the gas temperature and vapor pressure. Water vapor is generated continuously in the steam generator and injected in the closed loop through a control valve. The pressure is maintained slightly above atmospheric pressure using a pressure control valve.

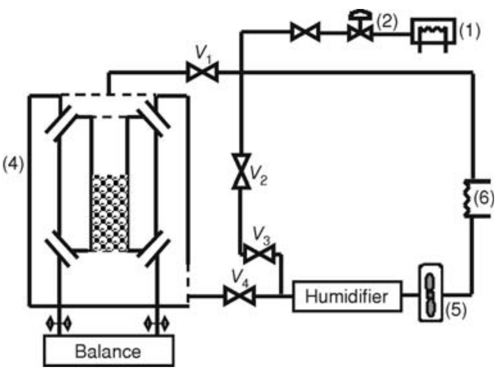

Figure 1. The experimental device
1– steam generator, 2 – pressure control valve, 3 – steam flow rate valve, 4 – convective cell, 5 – fan, 6 – hot gas heating system

The cell is an opaque aluminum cylinder. The inside diameter and the length of the cell were approximately 0.18 and 0.32 m, respectively.

Humid air flows through the cell containing the packed bed of alumina particles. At the inlet of the cell, a grid ensures a good distribution of humid air. Thermocouples of K-type are placed at different positions in the bed to measure the temperature of the alumina beads. During the packing of the bed in the experimental cell, thermocouples were introduced in the bed at the particle surface in different axial positions. Meanwhile, the experimental equipment was heated to a desired experimental temperature and the steam generator was activated. Steady-state conditions were reached. Then the experimental cell was inserted. The data acquisition system was started immediately after closing the cell door, with continuous measurement of the alumina temperatures at different depths in the bed. The experiment was ended when the alumina particles temperature approached that of the gas temperature. The characteristics of alumina particles are listed in tab. 1.

Table 1. Properties of alumina particles

Parameters	Alumina
d [mm]	4.35
ρ [kgm^{-3}]	1570
ε	0.4
C_{ps} [Jkg^{-1}K^{-1}]	840
λ_s [Wm^{-1}K^{-1}]	1

Computation of gas to particle heat transfer

The temperature of the particle surface that is necessary to quantify the heat transfer can be conveniently described in term of the gas to particle heat transfer coefficient. Based on experimental temperature measurements obtained in this work, a simple model was developed considering a steady-state energy balances and assuming (1) that the granular medium is homogeneous and isotropic, (2) that the bed porosity is uniform, (3) that the operation is adiabatic, and (4) that the bed is considered as a two-phase mixture (solid and gas) [13-16]. The packed bed is a discontinuous granular medium made of spherical porous particles (solid phase) crossed by humid air (gas phase). Energy balances are made in a bed portion whose height and cross-section are dz and S_0 (fig. 2), respectively.

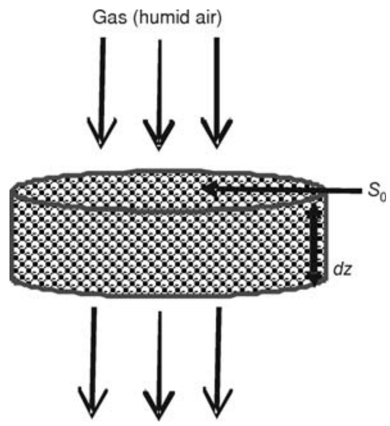

Figure 2. Packed bed section

The solid phase internal energy (U_s) is relayed to the evaporation rate $\dot{m}$ and the gas temperature:

$$\frac{\mathrm{d}U_{\mathrm{s}}}{\mathrm{d}t} = -\dot{m}C_{\mathrm{pg}}\frac{\mathrm{d}T_{\mathrm{g}}}{\mathrm{d}z}\mathrm{d}z \tag{1}$$

with C_{pg} and T_g are the gas specific heat and gas temperature, respectively.

For the gas phase the energy balance is:

$$\dot{m}C_{\mathrm{pg}}\frac{\mathrm{d}T_{\mathrm{g}}}{\mathrm{d}z} = h(T_{\mathrm{b}} - T_{\mathrm{g}})\frac{\mathrm{d}S_{\mathrm{b}}}{\mathrm{d}z} \tag{2}$$

where h is the gas to particle heat transfer and S_b – the total surface area of the particles.

For spherical particles:

$$S_{\mathrm{b}} = \frac{6(1-\varepsilon)S_{0z}}{d} \tag{3}$$

where ε is the porosity and d is the particle diameter.

By replacing eq. (1) by (2):

$$\frac{\mathrm{d}U_s}{\mathrm{d}t} = -h(T_{\mathrm{b}} - T_{\mathrm{g}})\frac{\mathrm{d}S_{\mathrm{b}}}{\mathrm{d}z}\mathrm{d}z \tag{4}$$

Integration results on:

$$\ln\left(\frac{T_{\mathrm{b}} - T_{\mathrm{g}}}{T_{\mathrm{b0}} - T_{\mathrm{g}}}\right) = -\frac{h(1-\varepsilon)S_0 H}{MC_{\mathrm{ps}}d} \tag{5}$$

where T_{b0}, M, H, and C_{ps} are initial bed temperature, bed weight, bed height, and solid specific heat, respectively.

This equation indicates that a plot of the logarithmic temperature $\ln(T_b - T_g)/(T_{b0} - T_g)$ *vs*. time should present a straight line with a slope equal to $h(1-\varepsilon)S_0H/MC_{ps}d$. The value of h can be given from this slope.

Results and discussion

Figure 3 shows that three temperatures measurements can be performed in the bed. At the entry of the cylinder ($z = 0$), the product reaches the gas temperature more quickly than the

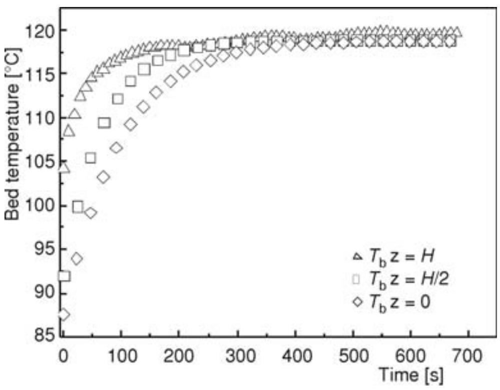

Figure 3. Variation of bed temperature with time; T_g = 120 °C, V_g = 2 m/s

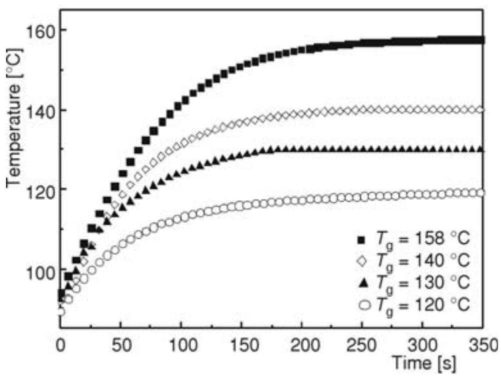

Figure 4. Variation of bed temperature with time; V_g = 2 m/s

center ($z = H/2$) and the exit ($z = H$). Next, the temperature presented in this study is at $z = H/2$.

Figures 4 and 5 show the bed temperature against time for different values of temperature and gas velocity, respectively. As the temperature increases, the bed reaches the gas temperature more quickly. This is because the higher temperature gradient in the medium resulted in a higher heat flux [17, 18]. As we would expect, when we increase the gas velocity, for the same

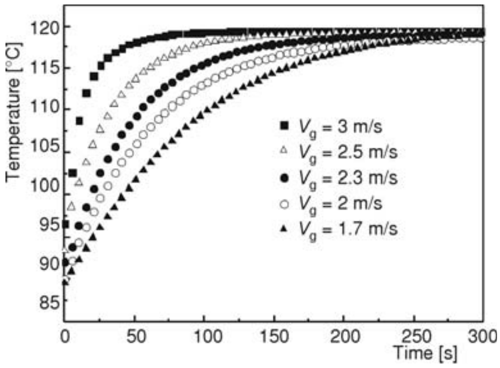

Figure 5. Variation of bed temperature with time: T_g = 120 °C

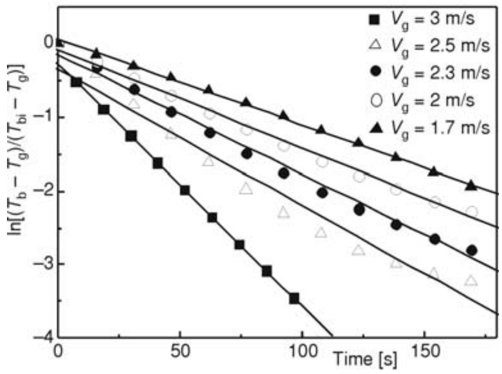

Figure 6. Straight lines obtained form eq. (5)

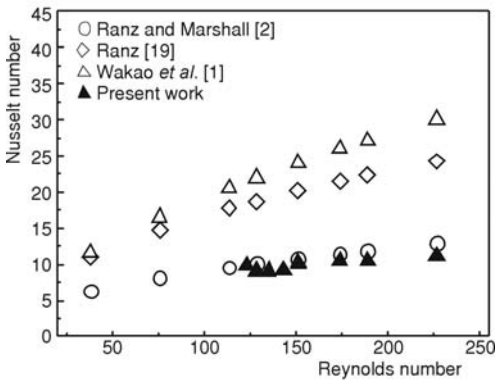

Figure 7. Nusselt *vs.* Reynolds numbers and compared to other works

gas temperature, the Reynolds number and consequently the convective heat transfer coefficient increases leading to a reduction of the time needed to reach the gas temperature.

Figure 6 shows an example of these logarithmic temperatures differences *vs.* time for different values of gas velocity. As expected, a family of straight lines was produced. The determination of slopes of these lines leads to the measured heat transfer coefficient and consequently the experimental Nusselt number. Hence, fig. 7 illustrates the relation between Nusselt and Reynolds numbers for the model developed in this study compared with correlations reported in the literature for packed bed of spherical porous alumina particles [1, 2, 19]. It can be seen that a general correlation relating Nusselt with Reynolds for different materials with a wide range of physical and transport properties is not possible. This has been generally observed in the literature where the correlation for the heat transfer coefficient of one author working with one materiel is often at variance with that of other author working with some other materials. In fact, the heat transfer coefficient is dependent both on the thermal properties of the bed and on the flux rate.

However, most experimental correlations express it as a function of the Reynolds number. Therefore, the applicability of these correlations is also limited to the particular bed materials used in developing them. Moreover, the thermal properties of the walls (adiabatic or heated) have a significant effect on the conduction mode and therefore studies using similar experimental techniques but different bed materials can also be expected to yield different correlation relating the heat transfer coefficient to Reynolds numbers. Tsotsas and Schlunder [20] showed that the non-uniform distribution of gas velocity, gas thermal conductivity, axial distribution of heat and errors in measurement caused discrepancies of wall heat transfer coefficient values. Yagi and Kunii [21] discussed the variation of heat transfer coefficient near the wall in the literature and proposed an equation to calculate this coefficient in cylindrical packed beds for low Reynolds numbers. Dixon *et al*. [22] explained the variation of apparent wall heat transfer coefficient values and showed the importance of predicting this coefficient.

We note also that the correlations of Nusselt number used to validate our study give a Re $\rightarrow$ 0 asymptote of Nusselt equal to 2, which is reasonable for spherical particles. We note also that at low Reynolds number, the interfacial convection heat transfer is insignificant compared to other terms in energy equation, and therefore, the suggested Re $\rightarrow$ 0 asymptote cannot be experimentally verified [23]. As explained above, discrepancies in Nusselt number were found. This may be due, also, to some differences in experimental conditions (particles characteristics, height and diameter of the packed bed, boundary conditions...). These discrepancies can be also attributed also to the position of the thermocouples and errors in the measurement [24, 25].

For these reasons, a simple model based on experimental measurements allowing the calculation of heat transfer coefficient for each packed bed material with moderate experimental instrumentations is developed in the present work.

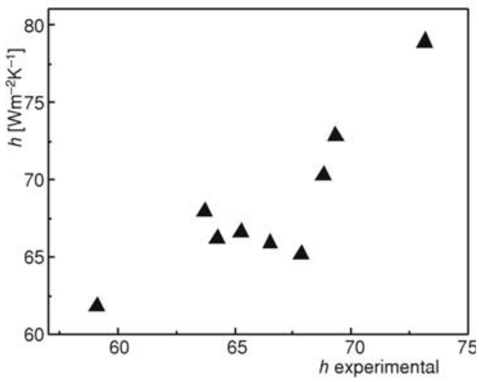

Figure 8. Heat transfer coefficient according to Ranz and Marshall correlation *vs.* experimental values

Conclusions

Due to the lack of a simple experimental procedure and equations for predicting the gas to particle heat transfer coefficient of packed beds for each material and experimental conditions especially at high temperature, an attempt is made to calculate this coefficient at low Reynolds numbers. The method is based on bed temperature measurements and simple energy balance. A logarithmic mean temperature difference method is used. Nusselt numbers obtained from the literature are represented *vs.* those calculated from the proposed model. Discrepancies were found. Consequently, the difference in experimental conditions used to establish these correlations modifies values of calculated Nusselt numbers. Next, the heat transfer coefficients obtained were compared to these in the literature.

A reasonable agreement was found with the widely used Ranz and Marshall correlation (fig. 8).

Thus, a needed and simple way to determine heat transfer coefficient was developed and verified. The model can be useful to predict gas to particle heat transfer coefficient in design and modeling of fixed beds for drying or combustion.

Acknowledgments

Nomenclature

C_p – specific heat, [Jkg^{-1}K^{-1}]
d – particle diameter, [m]
H – bed height, [m]
h – heat transfer coefficient at particle to gas interface, [Wm^{-2}K^{-1}]
M – bed weight, [kg]

$\dot{m}$	–	evaporation rate, [$kgm^{-3}s^{-1}$]
Nu	–	Nusselt number
P	–	pressure, [bar]
Pr	–	Prandtl number (= $C_{pg}\mu_g/\lambda_g$)
S_b	–	surface area of particles, [m^2]
S_0	–	bed cross-section, [m^2]
T	–	temperature, [°C]
t	–	time, [s]
Re	–	Reynolds number, (= $\rho_g V_g d/\mu g$)
U	–	internal energy, [J]
V	–	gas velocity, [ms^{-1}]
z	–	axial co-ordinate, [m]

Greek symbols

ε	–	porosity
μ	–	dynamic viscosity, [Pa·s]
λ	–	thermal conductivity, [$Wm^{-1}K^{-1}$]
ρ	–	density, [kgm^{-3}]

Subscripts

b	–	bed
g	–	gas
s	–	solid
v	–	vapor

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