On the Influence of Liquid Distribution on Heat Transfer Parameters in Trickle Bed Systems

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The aim of this work is to analyse experimental results on radial heat transfer in a packed bed column with cocurrent downflow of liquid and gas, focused on the influence of radial liquid distribution and column to particle diameter ratio. For this purpose, two kinds of measurements were carried out: radial liquid distribution and radial temperature profiles. Heat transfer data as analysed with a model consisting of two regions: a lumped wall zone up to a distance of a particle radius from the tube wall, and a distributed core region covering the rest of the bed. Three parameters arose: the radial effective thermal conductivity of the core region ($k_{er,c}$), the heat transfer coefficient between both zones (h_l), and the heat transfer coefficient from the wall region to the tube wall ($h_{w,w}$). Suitable correlations for the mentioned parameters as a function of gas and liquid Reynolds number and particle diameter were developed.

Ce travail a pour but d'analyser les résultats expérimentaux sur le transfert de chaleur radial dans une colonne à lit garni à écoulement descendant à cocourant de liquide et de gaz, en s'intéressant en particulier à l'influence de la distribution radiale de liquide et du rapport entre la colonne et le diamètre des particules. Pour ce faire, deux sortes de mesures ont été effectuées, soient les profils radiaux de distribution de liquide et les profils radiaux de température. Les données de transfert de chaleur ont été analysées à l'aide d'un modèle composé de deux régions : une zone de paroi moyennée sur une distance correspondant à un rayone de particule à partir de la paroi du tube et une région centrale représentant le reste du lit. Trois paramètres apparaissent : la conductivité thermique effective radiale de la région centrale ($k_{er,c}$), le coefficient de transfert de chaleur entre les deux zones (h_l) et le coefficient de transfert de chaleur entre la région de la paroi jusqu'à la paroi du tube ($h_{w,w}$). Des corrélations appropriées ont été établies pour les paramètres mentionnés en fonction du nombre de Reynolds du gaz et du liquide et du diamètre de particule.

Keywords: heat transfer parameters, two zones model, trickle bed reactors, wall effect, radial liquid distribution.

Design and analysis of multiphase reactors has become one of the most intensively studied topics in chemical reaction engineering. Moreover, the field of application of such type of reactors is increasing daily. In this context, trickle bed reactors (TBR) are frequently employed in many chemical, petrochemical and environmental applications. In spite of this, only a few articles in available literature focus on the influence of aspect ratio on thermal and fluid-dynamics behaviour of such systems.

Almost all the experimental studies dealing with heat transfer in TBR were carried out with aspect ratios (column to particle diameter ratio) greater than 10, and none of them include the influence of radial liquid distribution. However, there are many examples of industrial processes involving exothermic reactions, which are conducted in multitubular TBR (synthesis of MIBK -methyl-isobutyl ketone, or advance Fischer-Tropsch technology, to convert natural gas into synthetic hydrocarbon (Sie and Krishna, 1999). In such systems, particle diameter is of the same order of magnitude as tube diameter and low aspect ratios can certainly arise.

In general, the two-dimensional pseudohomogeneous plug flow (2DPPF) model with two parameters (uniform effective thermal conductivity, k_{er}, and wall heat transfer coefficient, h_w) was used to represent the heat transfer phenomena in TBR. A concise review on the subject can be found in Lamine et al. (1996).

Concerning our own experimental heat transfer measurements, the heat transfer process was analysed by means of a 2DPPF model in a first attempt (Mariani et al., 2001). k_{er} was well correlated with liquid and gas Reynolds numbers and particle diameter, except for the lowest experimental aspect ratio, $a = 4.7$ (Mariani et al., 2001).

Instead, the behaviour shown by the values of h_w was different. The values of Nusselt number, Nu_w, are plotted in Figure 1, where it can be observed that the effect of liquid Reynolds number, Re_L, on Nu_w is clearly different at high ($a = 34.3$, 17.2) and low ($a = 8.2$, 4.7) aspect ratios. The correlation included in Figure 1 was previously developed (Mariani et al., 2001), for only the high aspect ratios, $a = 34.3$, 17.2, as a single dependency on the whole range of Re_L can apparently be ruled out. A much

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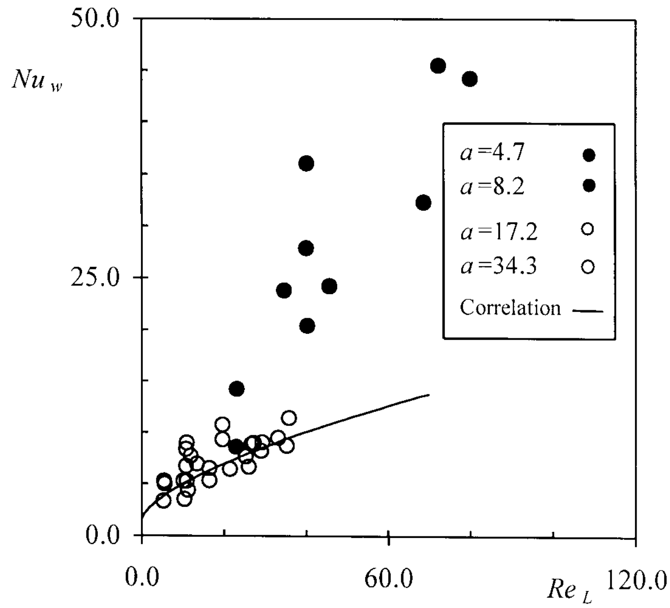

Figure 1. Nu_w of the 2DPPF model vs liquid Reynolds number.

stronger effect of Re_L on Nu_w is evident for the low aspect ratios.

The reason for this behavior may be found in the plug flow assumption of the model. It is well known that the zone from the wall up to around half particle diameter, the so called wall-zone, shows a larger void fraction than the rest of the bed (Mariani et al., 2002). A higher permeability is then expected in that zone, a feature that at least for one-phase flow has been clearly demonstrated (Dixon et al., 1984). While the amount of fluid flow in the wall-zone, noticeably liquid flow, can be justifiably ignored in the frame of the 2DPPF model at large values of a, the effect can be hardly neglected at low values: the wall-zone represents 5.7% of the total cross section area for a = 34.3 and 38.2% for a = 4.7. It is then reasonable to believe that the high values of Nu_w (Figure 1) resulting from applying the 2DPPF model arise mainly as a consequence of ignoring the extra amount of liquid flowing in the wall-region, which can be actually heated more easily than the assumption of uniform flow distribution allows.

The purpose of the present contribution is to report measurement of the radial liquid distribution in order to develop a model capable to take into account such distribution. In addition, this model will be employed to re-analyse the heat transfer data previously reported (Mariani et al., 2001).

Experimental

Heat Transfer Measurements

The experimental method used to study heat transfer between the bed and an external heating source is based on measuring the radial temperature profile at two angular positions and for different bed heights. The heat source was hot water circulating through a jacket surrounding the packed bed tube. The jacket was divided into three vertical zones, thus allowing different active heat transfer length. Glass spherical particles of four different sizes, d_p = 11.0, 6.3, 3.0, 1.5 mm, were packed in a 51.4 mm diameter tube. The corresponding aspect ratios were a = 4.7, 8.2, 17.2, 34.3. Water and air were employed as model fluids at normal atmospheric pressure and temperature. The superficial liquid flow rate was varied from 2.4 to 8.0 kg $m^{-2}{\cdot}s^{-1}$ and the gas flow rate from 3 10^{-2} to 0.5 kg $m^{-2}{\cdot}s^{-1}$. All the experiments were conducted in low interaction regime (trickle regime).

It is widely known that the fluid dynamic and thermal behavior of the trickle bed columns depend on the path followed to reach the steady state (Sundaresan, 1994). Therefore, it was highly convenient to set a standard experimental procedure to start up the experiments. The procedure includes fluidizing the bed with water which also guarantees a random packed bed structure for each experiment. In this context, every experimental condition (fixed gas and liquid flow rate, bed height and particle diameter) was replicated an average of six times. In each replication the temperature profile was recorded during 70 minutes summing up 500 temperature profiles.

Finally, 478 experiments were carried out including different bed heights, liquid and gas flow rate and particle sizes. Further details of the experimental set-up and procedure can be found in Mariani (2000), along with the range of variables for which measurements were performed.

Radial Liquid Distribution Measurements

Measurements of radial liquid distribution were conducted in a 1.3 m high Plexiglas column with 47.8 mm inner diameter. The largest particles used for the heat transfer tests, namely d_p = 11 mm, were employed. The resulting aspect ratio a = 4.3 was then close to the lowest value for the heat transfer experiments (a = 4.7). The fluids were again air and water at room temperature. Liquid and gas flow rates were varied within the range allowed by the low interaction regime (trickle regime). Two distributor plates were used to check the effect of inlet liquid distribution: one with five 2.5 mm holes and the other with thirty 1 mm holes (I.D.).

The liquid collector at the bottom of the bed was divided into two concentric regions. One of them collected the liquid from the bed wall up to half particle diameter (wall region) and the other collected the liquid from the rest of the bed cross section (core region). This definition of the wall region relies on the physical evidence that there is minimum value of porosity at a distance of about a half particle diameter from the wall (Mariani et al., 2002). Hence, this distance naturally defines two different zones of the bed.

The liquid flow rate in each region, $q_{L,w}$ and $q_{L,c}$ was calculated by measuring the time elapsed to collect a given volume.

The start up procedure and the system employed to control gas and liquid flow rates were the same as for the heat transfer experiments (Mariani, 2000). Every experimental condition was also replicated an average of six times.

Experiments covered 78 different settings resulting from varying gas and liquid flow rates restricted to trickle regime.
The observed values of $q_{L,w}/q_L$ turned out to be statistically independent of the liquid and gas overall flow rates and of the type of distributor plate, showing an average value of approximately 80%. From this value, the average superficial velocity ratio between wall and core regions is $R_u = u_{L,w} / u_{L,c} \cong 6$.

It is worth recalling that for a single phase flow the ratio R_u takes, at most, values around 2 (Papageorgiou and Froment, 1995). The major difference with the present system should be found in distinct hydrodynamics features. For a single phase flow, the fluid that fills all the bed voids and the flow distribution between both regions will depend on the intrinsic permeability in both regions under a uniform pressure gradient throughout the cross section. In a trickle bed, the liquid phase is far from filling the bed voids and it essentially flows by the effect of gravity forming a film either on the particle surfaces or on the wall surface. The flow distribution between both regions will depend on the intrinsic resistances offered by the surfaces, but also on the film thicknesses. There is no restriction fixing the relative width of the thicknesses in both regions (i.e. we can envisage a situation presenting 100% of the liquid flowing at the wall if the liquid load is low enough to avoid flooding of the wall region).

In the course of the experiments, we were able to visualize that a nearly continuous film is developed on the wall surface.

Two-region Model for Heat Transfer Analysis

Following the approach presented for a single phase flow by Borkink and Westerterp (1994), among others, the packed bed is divided into two regions, each with a different but constant superficial velocity.

The two regions are a distributed core region and a lumped wall zone. A third lumped region is also included, that accounts for the exchanging fluid flow within the bed jacket. A sketch of the model is shown in Figure 2.

The wall region width was defined according to the description devised in the Introduction and Experimental sections.

A heat balance on a volume element leads, in the absence of any heat source, to:

Region c (distributed core region) $$w_c \frac{\partial T_c}{\partial z} = \lambda \frac{\partial(\rho \partial T_c / \partial \rho)}{\rho \partial \rho} \quad (1.1)$$

Region w (lumped wall region) $$w_w \frac{dT_w}{dz} = H_w(T_J - T_w) - H_l(T_w - T_c(1)) \quad (1.2)$$

Region J (jacket region - auxiliary fluid) $$w_J \frac{dT_J}{dz} = -H_w(T_J - T_w) \quad (1.3)$$

Boundary conditions:

Interface between regions w and c $$(\partial T_c / \partial \rho)\big|_{\rho=1} = -Bi(T_c(1) - T_w) \quad (1.4)$$

Symmetry condition $$(\partial T_c / \partial \rho)\big|_{\rho=0} = 0 \quad (1.5)$$

Inlet conditions $$z = 0: \quad T_c = T_{c0}; \quad T_w = T_{w0}; \quad T_j = T_{j0} \quad (1.6)$$

where:

$w_i = q_{G,i}\, C_{pG,i} + q_{L,i}\, C_{pL,i}$ $\quad (i = c, w, J)$;

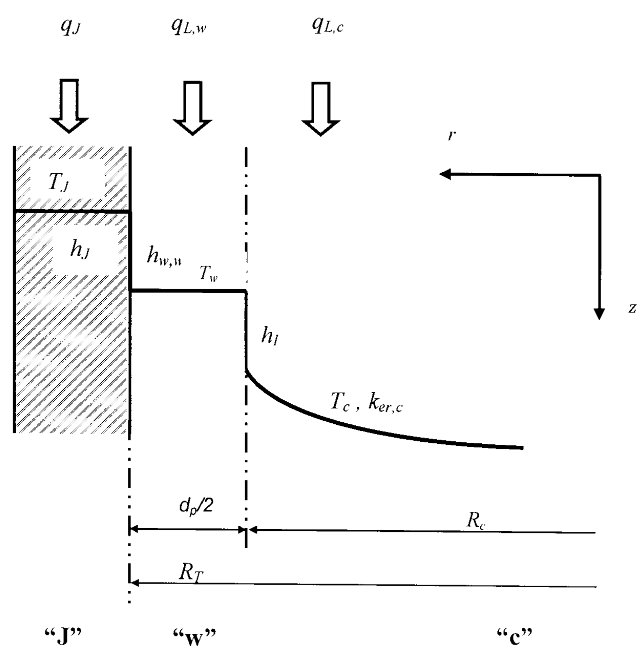

Figure 2. Sketch showing the two region model of the bed and exchanging jacket.

$H_w = 2\pi R_T (1/h_{w,w} + 1/h_J)^{-1}$ $\qquad H_l = 2\pi R_c\, h_l$;
$Bi = R_c\, h_l / k_{er,c}$; $\qquad \lambda = \pi k_{er,c} = H_l / (2\ Bi)$

The solution of Equation (1) is,

$$\begin{bmatrix} T_c - T_\infty \\ T_w - T_\infty \\ T_J - T_\infty \end{bmatrix} = \sum_{i=1}^{\infty} c_i \exp\left(-z \frac{\lambda \beta_i^2}{w_c}\right) \begin{bmatrix} \theta_i \\ \varphi_i \\ \eta_i \end{bmatrix} \quad (2)$$

where:

$T_\circ = (w_J T_{J0} + w_w T_{w0} + w_c \overline{T}_{c0})/(w_J + w_w + w_c)$

β_i are the eigenvalues resulting from solving the equation:

$\beta J_1(\beta) P(\beta) - Bi J_0(\beta) Q(\beta) = 0$

with $P(\beta) = 1 - Dx - x(1 - Ax)$; $\quad Q(\beta) = 1 - Dx$;

$x = (2\ Bi\ w_c / w_w)/\beta^2$; $\qquad A = (H_w/H_l)(w_w/w_J)$;
$D = (H_w/H_l)(1 + w_w/w_J)$

θ_i, are the eigenfuntions of the distributed core region and φ_i, η_i are the eigenconstants of wall and jacket regions:

$\theta_i = J_0(\beta_i\, \rho)$
$\varphi_i = J_0(\beta_i) - \beta_i\, J_1(\beta_i)/Bi$
$\eta_i = [\beta_i\, J_1(\beta_i)(w_w/Bi - 2w_c/\beta_i^2) - w_w J_0(\beta_i)] / w_J$

The coefficients c_i can be determined using the orthogonality relation

$$2w_c\int_0^1 \theta_m\theta_n\rho d\rho + w_w\varphi_m\varphi_n + w_J\eta_m\eta_n = 0, \text{ if } m \neq n \quad (3)$$

Then,

$$c_i \equiv c_n\begin{bmatrix} 2w_c\int_0^1 (T_c - T_{c0})\theta_n\rho d\rho + w_w(T_w - T_{w0}) \\ \varphi_n + w_J(T_J - T_{J0})\eta_n \end{bmatrix} / \mathrm{LOC}(n,n) \quad (4)$$

where: $\mathrm{LOC}(n, n) = w_c[J_0^2(\beta_n) + J_1^2(\beta_n)] + w_w\,\varphi_n^2 + w_J\eta_n^2$

Details about deriving Equation (2) from Equation (1) can be requested from the corresponding author.

The heat transfer coefficient h_J between the jacket fluid and the wall surface was evaluated as previously described (Mariani, 2000): $h_J = 10632\ \mathrm{W \cdot m^{-2} \cdot {}^\circ C^{-1}}$.

Identification of the Parameters of the Two-region Model

The two-region model described above introduces three thermal parameters: the radial effective thermal conductivity of the core region ($k_{er,c}$), the heat transfer coefficient between both regions (h_I), and the heat transfer coefficient from the wall region to the tube wall ($h_{w,w}$). It is intended in this section to infer the values of these parameters on the basis of the experimental data from the heat transfer experiments outlined above.

To this end, it is first necessary to assign values to the liquid flow rates in each section $q_{L,w}$ and $q_{L,c}$. We have assumed $R_u=u_{L,w} / u_{L,c} = 6$ for the four particle sizes employed in the experiments. From this ratio, the values $q_{L,w}$ and $q_{L,c}$ follow immediately from the known value of the overall liquid flow rate q_L in each test. The ratio $R_u = 6$ was determined for the 11 mm particles and under a similar aspect ratio a as the one resulting in the heat transfer experiments with these particles. Therefore, that assignment is most likely to be correct in this case. There is an increasing uncertainty whether the value $R_u = 6$ will hold as d_p decreases, but it is simultaneously true that the effect of R_u will be less and less significant. In fact, at the lowest aspect ratio ($a = 4.7$, $d_p = 11$mm) the flow at the wall region with $R_u = 6$ is 78.8%. Therefore, most of the liquid input exchanges heat from the wall region. At the highest aspect ratio ($a = 34.3$, $d_p = 1.5$ mm), only 26.8% of the liquid input flows in the wall region when $R_u = 6$. This is a low fraction that will remain so even if some variation of R_u is allowed. What is most important, with respect to the purpose of this work is to evaluate the sensibility of the heat transfer parameters to the value of R_u. It has been checked that lowering R_u in 25% produces changes in the best estimates of h_I and $k_{er,c}$ of about 7%. This variation falls within the limits of confidence discussed below.

In order to strictly define the flowing heat capacity in both region, $w_w = q_{G,w}\,C_{pG,w} + q_{L,w}\,C_{pL,w}$ and $w_c = q_{G,c}\,C_{pG,c} + q_{L,c}\,C_{pL,c}$, the contribution from the gas flow has to be accounted for. Taking into account that this is a minor contribution, a uniform gas flow distribution was assumed to calculate w_w and w_c.

When trying to adjust the three thermal parameters $h_{w,w}$, h_I and $k_{er,c}$ from a regression analysis of the experimental data, a very strong correlation between $h_{w,w}$ and h_I was found. This result is most likely due to the distribution of the set of thermocouples close to the bed exit being fixed for the different aspect ratios. In this way, the temperature difference close to boundary between both zones could not be properly evaluated from the experimental readings. Therefore, the parameter $h_{w,w}$ was independently estimated and the values $k_{er,c}$ and h_I were obtained from regression analysis as follows:

$h_{w,w}$ will depend directly on the thickness δ of the film falling just on the wall surface (the larger δ, the lower $h_{w,w}$). To evaluate δ, it was assumed that the whole liquid flow in the wall region takes place just on the wall surface. The highest possible value of δ will then result. The expression for a laminar film on a flat surface was employed (Bird et al., 2001):

$$\delta = \sqrt[3]{\frac{3\mu\Gamma}{\rho_L^2 g}} \quad (5)$$

where $\Gamma = q_{L,w}/(2\pi R_T)$ is the liquid loading rate and $q_{L,w}$ is the value obtained from the experimental liquid distribution results.

The corresponding value of heat transfer coefficient between the film and the wall surface (Bird et al., 2001) was then employed to estimate $h_{w,w}$:

$$h_{w,w} = 2\frac{k_L}{\delta} \quad (6)$$

An average film temperature of 60°C was assumed to evaluate physical and transport properties in the film. The resulting values of δ lay in the range 0.1 to 0.22 mm and it was checked that effectively the laminar regime took place. Those values of δ are about one tenth of the smallest particle diameter, suggesting that ignoring particle surfaces in the wall region for estimating δ could not be much significant.

The final merit in estimating $h_{w,w}$ can be assessed by comparing their values with the values of h_I obtained from regression of the experimental data. Values of $h_{w,w}$ were in the range 6300-13000 $\mathrm{W \cdot m^{-2} \cdot C^{-1}}$ and were at least five times the corresponding value of h_I (see section Heat transfer coefficient between regions). A sensitivity test was also made by fitting h_I with values of $h_{w,w}$ varied in 25%. On average, the variation of h_I was only 8%. These results indicate that the approach followed by estimating $h_{w,w}$ independently was satisfactory.

A typical radial temperature profile obtained in the experiments is shown in Figure 3; it is also included the profile predicted by the model presented in this contribution.

Effective Thermal Conductivity in Core Region

To finally estimate $k_{er,c}$ and h_I, both were considered to be independent of the heating length. The Greg Software Package (Stewart et al., 1992) was employed for the regression analysis.

The behaviour of $k_{er,c}$ was completely similar to that of the analogous parameter k_e in the 2DPPF model (Mariani et al., 2001). The gas Reynolds number, Re_G, shows a weak but definite influence on the fitted values of $k_{er,c}$. An increase of about 30% in $k_{er,c}$ was observed from the lowest to the highest values of Re_G. However, the liquid phase plays the most important role over the range of experimental conditions. $k_{er,c}$ is expressed by adding the contributions from the bed without fluid-flow and from the lateral mixing:

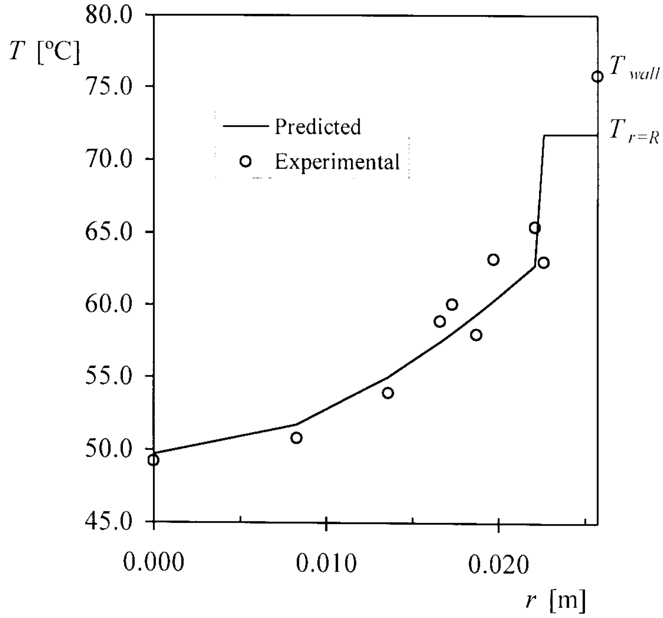

Figure 3. Radial temperature profiles. G_L = 7.7 kg·m^{-2}·s^{-2}, G_G = 4 10^{-2} kg·m^{-2}·s^{-1}, d_p = 6.3 mm.

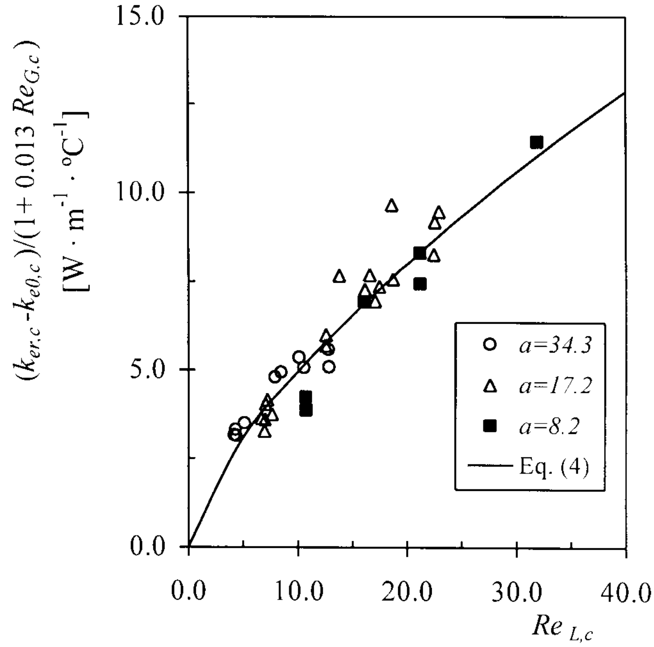

Figure 4. Experimental and correlated values of $k_{er,c}$ vs $Re_{L,c}$.

$$k_{er,c} = k_{e0,c} + ak_L(1 + bRe_G)Re_{L,c}^d Pr_L \tag{7}$$

where $Re_{L,c}$ is the liquid Reynolds based on the core superficial velocity $u_{L,c}$, a, b and d were fitting parameters and $k_{e0,c}$ is the contribution without fluid-flow, estimated as detailed by Mariani (2000). This contribution was at least one order of magnitude lower than the flow contribution. The term (1 + bRe_G), with $b > 0$, represents an enhancement effect of the gas-flow to liquid-flow lateralization.

After fitting all data, except those for d_p = 11 mm (a = 4.7), Equation (7) becomes:

$$k_{er,c} = k_{e0,c} + 0.36k_L(1 + 0.013Re_G)Re_{L,c}^{0.69} Pr_L \tag{8}$$

Experimental data and values from Equation (8) are given in Figure 4 as a function of $Re_{L,c}$. The average deviation, defined as

$$\Delta_k = \frac{1}{N}\sum_{i=1}^{N}\left|\frac{k_{er,c}^{obs} - k_{er,c}^{corr}}{k_{er,c}^{obs}}\right|_i \tag{9}$$

where N is the number of experimental data points, is less than 8%. It should be noted that this deviation falls between the average confidence limits of $k_{er,c}$. Excluding the results from 11 mm particles, the confidence limits for $k_{er,c}$ were about ±11%. Instead, for 11 mm particles the confidence limits were in many cases of the same order of the own estimate $k_{er,c}$. This result can be ascribed to the relatively smooth temperature profiles found at the lowest values of liquid flow rates for that particle size.

Heat Transfer Coefficient Between Regions

The results for h_l were not significantly sensitive to the gas flow rate. On the other hand, under the present experimental conditions, the liquid flow rate presents a strong effect on h_l. This results suggests a dependency of $Nu_l = h_l\, d_p / k_L$ on $Re_{L,c}$.

The fitted values of Nu_l for the whole set of particles as function of $Re_{L,c}$ are plotted in Figure 5. The behaviour of Nu_l was correlated with operating conditions by means of the following expression:

$$Nu_l = 0.36Re_{L,c}^{0.76}Pr_L^{1/3} \tag{10}$$

In the Equation (10), the usual Chilton-Colbun dependency of Nu on Pr was assumed.

It can be clearly appreciated from Figure 5 that the values of Nu_l for all particle diameters follows a continuous trend as $Re_{L,c}$ increases. It is also evident that Equation (10) is able to correlate satisfactorily the results. The average deviation, defined as:

$$\Delta_{Nu} = \frac{1}{N}\sum_{i=1}^{N}\left|\frac{Nu_l^{obs} - Nu_l^{corr}}{Nu_l^{obs}}\right|_i$$

is less than 18%. In this case, the confidence limits for Nu_l were around ±12%.

A comment should be made about the actual mechanisms leading to define the coefficient h_l. In principle, a contribution from lateral exchange between wall and core region can be expected. Nonetheless, this lateralization should be restrained by the small size of the open windows left among the particles in the layer adjacent to the wall. It can be envisaged that an additional parallel mechanism can exist, by considering heat exchange between fluid in the wall region and particles, in series with conduction through the particles followed by

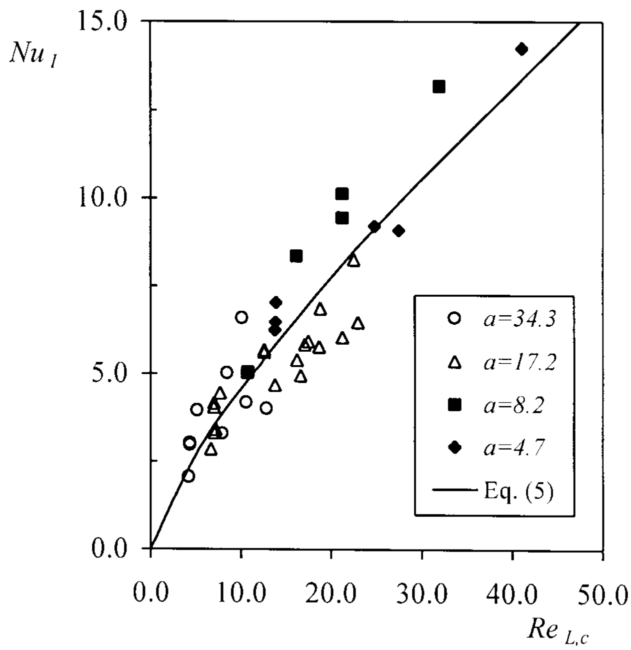

Figure 5. Experimental and correlated values of Nu_l vs $Re_{L,c}$.

exchange between particles and fluid in the core region. At this stage, however, there is not enough quantitative information to develop a mechanistic model to describe the behaviour of h_l.

Conclusions

A previous analysis (Mariani et al., 2001) has revealed that the 2DPPF model is not appropriate to interpret heat transfer in trickle beds with low aspect ratios, a fact that has already been pointed out for one-phase flow. The increased flow in the region nearby the bed wall was strongly suspected to be the key factor for the limitations of 2DPPF model.

Attending to this situation, the results from an experimental investigation on heat transfer from a packed bed with cocurrent gas-liquid downflow to the wall have been analysed to include the mentioned effect. The set of data covers the range of operating liquid and gas flow rates of the trickle (low interaction) regime in beds presenting aspect ratios (tube to particle diameter ratio) from 4.7 to 34.3.

To quantify the wall effect, experimental measurements on liquid distribution have been carried out with an aspect ratio similar as the smallest employed in the heat transfer measurements. The results clearly show that the liquid distribution is highly non-uniform: the average ratio between wall and core superficial velocities was 6.

With this information, a two-region model with different velocities in the core and wall regions was developed to represent heat transfer from the trickle bed to the wall surface. This model introduces three thermal parameters: radial effective thermal conductivity of the core region ($k_{er,c}$), heat transfer coefficient between both zones (h_l), and heat transfer coefficient between the wall region and the tube wall ($h_{w,w}$). Values of $h_{w,w}$ were independently calculated by assuming that a liquid film flows on the tube wall. It was shown that this assumption is not likely to introduce much effect on the estimated values of the remaining parameters, $k_{er,c}$ and h_l.

Values of $k_{er,c}$ and h_l were fitted to the experimental data set. A couple of expressions to evaluate $k_{er,c}$ and Nu_l (Equation 4 and 5) in terms of operating variables were also developed. The performance of these expressions is good and they can be useful to predict the thermal parameters in the whole range of aspect ratios studied.

The effect of the gas flow rate deserves a general comment. Within the range of values employed in the experiments (low interaction regime), this variable shows little effect on the overall heat transfer rate or on the thermal parameters. The main reasons for this behaviour are the very low mass-flow contribution of the gas phase and the low interaction feature of the trickle regime.

Experiments with other particle shapes (non spherical) and with different liquid properties would be desirable to complete a body of basic experimental data. In addition, a mechanistic model for the heat transfer coefficient between regions is still open to further development.

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Nomenclature

a	bed to particle diameter ratio
R_u	wall to core superficial velocity ratio
Bi	Biot number, $R_c\, h_l / k_{er,c}$
C_{Pi}	specific heat of the i-th stream, $(J \cdot kg^{-1} \cdot °C^{-1})$
d_p	particle diameter, (m)
g	acceleration of gravity, $(m \cdot s^{-2})$
G_i	superficial mass flow rate of the i-th stream, $(kg \cdot m^{-2} \cdot s^{-1})$
h_w	wall heat transfer coefficient -2DPPF model, $(W \cdot m^{-2} \cdot °C^{-1})$
h_J	jacket heat transfer coefficient, $(W \cdot m^{-2} \cdot °C^{-1})$
h_l	heat transfer coefficient between wall and core regions, $(W \cdot m^{-2} \cdot °C^{-1})$
$h_{w,w}$	heat transfer coefficient between wall region and tube wall, $(W \cdot m^{-2} \cdot °C^{-1})$
k	fluid thermal conductivity, $(W \cdot m^{-1} \cdot °C^{-1})$
k_{er}	effective radial thermal conductivity -2DPPF model, $(W \cdot m^{-1} \cdot °C^{-1})$
$k_{er,c}$	effective radial thermal conductivity of the core region, $(W \cdot m^{-1} \cdot °C^{-1})$
q_i	mass flow rate of the i-th stream, $(kg \cdot s^{-1})$
Nu_w	Nusselt number at the wall -2DPPF model, $h_w\, d_p / k_L$
Nu_l	Nusselt number between wall and core regions, $h_l\, d_p / k_L$
Pr	Prandtl number, $C_P\, \mu / k$
Re	Reynolds number, $G\, d_p / \mu$
R_T	tube radius, (m)
R_c	radius of the core region, (m)
T_c	core region temperature, (°C)
T_w	wall region temperature, (°C)
T_J	jacket temperature, (°C)
u	velocity, $(m \cdot s^{-1})$

Greek Symbols

δ	liquid film tickness, (m)
Γ	liquid loading rate, $(kg \cdot m^{-1} \cdot s^{-1})$
ρ	dimensionless radius, r/R_c
ρ_L	liquid density, $(kg \cdot m^{-3})$
μ	dynamic viscosity, (Pa·s)

Sub scripts

G	gas
L	liquid
0	inlet

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