Thermal Resistance Models for Effective Heat Transfer Parameters in Trickle Bed Reactors

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ABSTRACT

The predictive formulae for estimating the heat transfer parameters (k_{er} – the effective radial thermal conductivity of the bed; h_w – the wall-to-bed heat transfer coefficient) for two cases (High Reynolds Number Case and Low Reynolds Number Case) are derived for two-phase gas-liquid flow through paced bed reactors (trickle bed reactors) using the thermal resistance networks. Experiments are carried out to obtain the heat transfer parameters k_{er} and h_w for co-current down flow of air and water through packed beds over a wide range of flow rates of air (0.01-0.898 kg/m^{2}s) and water (3.16-71.05 kg/m^{2}s) covering trickle, pulse, and dispersed bubble flow regimes in a 50 mm i.d. column employing ceramic spheres (2 mm), glass spheres (4.05 and 6.75 mm) and ceramic raschig rings (4x4 and 6.75x6.75 mm) as the packing materials. The predicted values of k_{er} and h_w from the proposed thermal resistance models for two-phase flow agreed well with those obtained experimentally.

KEYWORDS: trickle bed reactor; packed bed; thermal resistance models; effective radial thermal conductivity of the bed; wall-to-bed heat transfer coefficient

INTRODUCTION

The modeling of heat transfer is of great importance to predict, *a priori*, the heat transfer parameters in the analysis and design of industrial trickle bed reactors (two-phase gas-liquid flow through packed bed reactors). However, no rigorous theoretical modeling of heat transfer in trickle bed reactors is attempted due to the complexity of flow behavior. A few semi-theoretical models for single-phase flow[1,2] were reported which have been later extended to two-phase flow.[3,4,5] These models do not help in predicting the effective heat transfer parameters (k_{er} – the effective radial thermal conductivity of the bed; h_w – the wall-to-bed heat transfer coefficient), *in priori*, as the parameters involved in the model equations are to be estimated either from the experiments or from the reported correlations,[4,6] which may be valid only for the range of experimental conditions investigated. The heat transfer phenomenon in packed beds is generally described by one- or two-dimensional models. The two-dimensional pseudo-homogeneous two-parameter model is widely employed using experimental radial temperature profile to estimate the effective heat transfer parameters in packed beds.[7]

Simple predictive models, *a priori*, would be desirable to estimate the effective heat transfer parameters both in single- and two-phase flow through packed beds. Dixon (1985) derived predictive formulae for the effective heat transfer parameters by employing simple thermal resistance models for single-phase packed bed heat transfer.[8] The predictive formulae require the individual solid and fluid phase heat transfer parameters. These individual phase parameters are fundamental in nature which presumably may be readily correlated and extrapolated. In view of this, a simple predictive model which relates the effective heat transfer parameters, either in single-phase or in two-phase flow through packed beds with the individual solid and fluid phase parameters is useful and reliable even in the situations where extrapolation from laboratory conditions to industrial scale is required. The thermal resistance models proposed by Dixon (1985) for single-phase flow heat transfer in packed bed are a pointer in this direction.[8] However, no theoretical work has been reported for two-phase gas-liquid flow through packed bed. The present work is aimed at (i) broadening the scope of the thermal resistance models to two-phase flow through packed beds, and (ii) comparing the experimental data on heat transfer parameters (data obtained in the present study and the data reported in the literature) for two-phase gas-liquid down flow through packed beds with those predicted by the thermal resistance models.

THERMAL RESISTANCE MODELS FOR TWO-PHASE FLOW PACKED BED HEAT TRANSFER

The thermal resistance network for radial heat transfer for single-phase flow through packed bed is represented in terms of either a single resistance of the bed (Fig. 1a) or as the bed center resistance and hear wall resistance in series (Fig. 1b), and can be related as given by Eq. 1 and Eq. 2 respectively.[8]

$$R_{\mathrm{tot}} = \frac{1}{h_{\mathrm{m}}} = \frac{R}{4\bar{k}_{\mathrm{er}}} \qquad (1)$$

$$R_{\mathrm{tot}} = R_{\mathrm{c}} + R_{\mathrm{w}} = \frac{R}{4k_{\mathrm{er}}} + \frac{1}{h_{\mathrm{w}}} \qquad (2)$$

Eq. 1 represents a single parameter model, in which the thermal resistance of the bed is related either to an overall heat transfer coefficient, h_{m}, or to an average effective radial thermal conductivity of the bed, k_{er}. Eq. 2 represents a two-parameter model in which the thermal resistance of the bed is considered to be the sum of the bed center resistance and the resistance near the wall, and is related to the effective radial thermal conductivity of the bed center (k_{er}) and the wall-to-bed heat transfer coefficient (h_{w}). Dixon (1985) derived the predictive formulae for k_{er} and h_{w} in single-phase packed bed heat transfer by considering the solid phase and the fluid phase resistances in parallel for three cases: (i) High Reynolds number case (high *Re* case: fluid phase dominant), (ii) Low Reynolds number case (low *Re* case: solid phase dominant), and (iii) Intermediate Reynolds number case (Intermediate *Re* case).[8] Interphase resistance was neglected as its contribution to the effective radial thermal conductivity of the bed center was found to be insignificant. The k_{er} values predicted by his model equations for high Reynolds number case (fluid dominant) agreed well with the literature data of Kunii et al. (1968), Valstar et al. (1975), Dixon et al. (1978), and some of their unpublished data over a range of operating conditions $5.0 \leq D_t/d_p \leq 11.1$, and $60 \leq \text{Re} \leq 700$.[9,10,11]

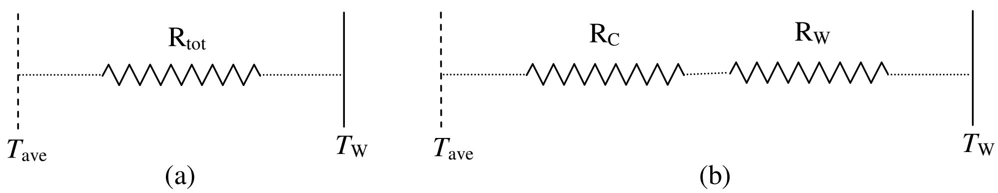

Figure 1. Thermal Resistance models of packed bed heat transfer.[8] (Dixon, 1985).

In view of a good agreement of the reported experimental data with the thermal resistance model predictions for single-phase flow, the thermal resistance networks[8] are extended for two-phase flow packed bed heat transfer by considering three resistances for gas, liquid, and solid. The liquid and gas phase resistances may be considered either in parallel or in series with the solid phase resistance. Yagi and Kunii (1957) developed a theoretical model for the effective bed radial thermal conductivity for single-phase flow through packed beds starting from the basic heat transfer mechanisms, considering the effective bed radial thermal conductivity as the summation of the radial thermal conductivity of the solid and the radial thermal conductivity due to lateral mixing of the fluid.[2] Later, this model was successfully extended to two-phase flow for both the effective radial thermal conductivity of the bed center and the wall-to-bed heat transfer coefficient, by adding the lateral mixing contributions of both the flowing phases to the solid contribution.[3,4,12,13,14] Based on these studies, the resistances of solid, gas and liquid can be considered to be in parallel. The schematic diagram representing these resistances is shown in Fig. 2a. For each phase, the resistance can be divided into the bed center and wall resistances which are in series, as shown in Fig. 2b. The resistance of the bed as represented by Figs. 2a and 2b is expressed as follows:

$$\frac{1}{R_{\text{tot}}} = \frac{1}{R_{\text{L}}} + \frac{1}{R_{\text{G}}} + \frac{1}{R_{\text{S}}} \tag{3}$$

where,

$$R_{\text{L}} = \frac{R}{4\bar{k}_{\text{rL}}} \equiv R_{\text{cL}} + R_{\text{wL}} = \frac{R}{4k_{\text{rL}}} + \frac{1}{h_{\text{wL}}} \tag{4}$$

$$R_{\text{G}} = \frac{R}{4\bar{k}_{\text{rG}}} \equiv R_{\text{cG}} + R_{\text{wG}} = \frac{R}{4k_{\text{rG}}} + \frac{1}{h_{\text{wG}}} \tag{5}$$

$$R_{\text{S}} = \frac{R}{4\bar{k}_{\text{rS}}} \equiv R_{\text{cS}} + R_{\text{wS}} = \frac{R}{4k_{\text{rS}}} + \frac{1}{h_{\text{wS}}} \tag{6}$$

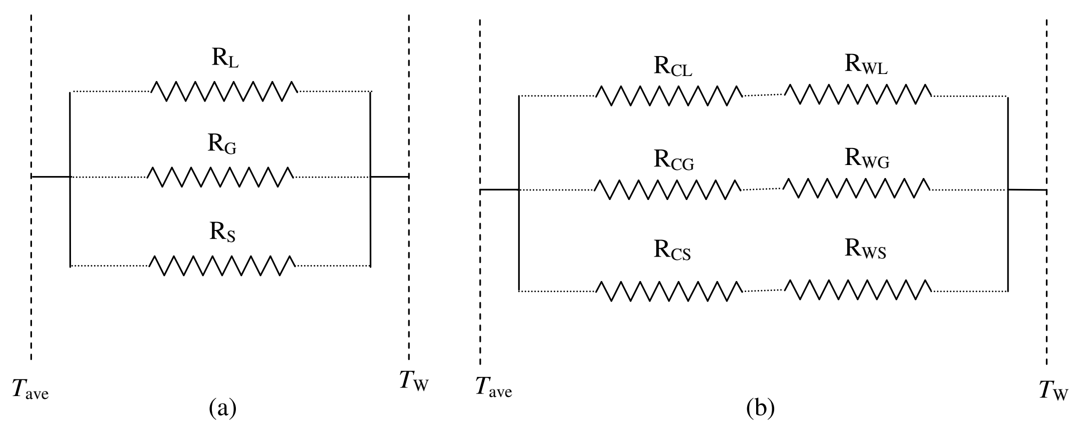

Figure 2. Thermal Resistance models of packed bed heat transfer for two-phase flow.

Using one-parameter model (Eq. 1), Eq. 3 can be written by relating the average effective radial thermal conductivity of the bed to those of the individual phases as:

$$\frac{4\bar{k}_{\text{er}}}{R} = \frac{4\bar{k}_{\text{rL}}}{R} + \frac{4\bar{k}_{\text{rG}}}{R} + \frac{4\bar{k}_{\text{rS}}}{R} \tag{7}$$

From Eqs. 1 and 2, $\bar{k}_{\text{er}}$ is related to Biot number, Bi (= h_{w}R/k_{er}), and k_{er}, as given by Eq. 8.

$$\bar{k}_{\mathrm{er}} = k_{\mathrm{er}}\left[\frac{Bi}{Bi+4}\right] \quad (8)$$

Similar equations can be obtained for the individual phases (Eqs. 9, 10, and 11 for liquid, gas, and solid phases).

$$\bar{k}_{\mathrm{rL}} = k_{\mathrm{rL}}\left[\frac{Bi_{\mathrm{L}}}{Bi_{\mathrm{L}}+4}\right] \quad (9)$$

$$\bar{k}_{\mathrm{rG}} = k_{\mathrm{rG}}\left[\frac{Bi_{\mathrm{G}}}{Bi_{\mathrm{G}}+4}\right] \quad (10)$$

$$\bar{k}_{\mathrm{rS}} = k_{\mathrm{rS}}\left[\frac{Bi_{\mathrm{S}}}{Bi_{\mathrm{S}}+4}\right] \quad (11)$$

Substituting Eqs. 8-11 in Eq. 7, and dividing by k_{L}, we get

$$\frac{k_{\mathrm{er}}}{k_{\mathrm{L}}}\left[\frac{Bi}{Bi+4}\right] = \frac{k_{\mathrm{rL}}}{k_{\mathrm{L}}}\left[\frac{Bi_{\mathrm{L}}}{Bi_{\mathrm{L}}+4}\right] + \frac{k_{\mathrm{rG}}}{k_{\mathrm{L}}}\left[\frac{Bi_{\mathrm{G}}}{Bi_{\mathrm{G}}+4}\right] + \frac{k_{\mathrm{rS}}}{k_{\mathrm{L}}}\left[\frac{Bi_{\mathrm{S}}}{Bi_{\mathrm{S}}+4}\right] \quad (12)$$

Eq. 12 contains two unknown quantities k_{er} and Bi. Dixon (1985),[8] in his analysis, used $Bi = Bi_{\mathrm{F}}$ (high Re case) as the second relation, which is applicable for single-phase packed bed heat transfer, to solve for k_{er} and Bi. This simplified assumption is not applicable for two-phase flow due to the presence of two fluid phases. Therefore, a second relation between Bi and the heat transfer parameters of individual phases of solid, gas, and liquid is obtained based on the following considerations.

The pseudo-homogeneous model for packed bed heat transfer is derived assuming that the temperature in the bed at any given radial position is same for all the phases. This implies that the radial temperature drop in each of the phases in the wall region is the same and so as at the bed center as well. Therefore, the thermal resistances of the individual phases can be taken as parallel to each other both at the center and near the wall. Hence, the ratio of bed center resistance to the wall resistance can be written as shown in Eq. 13.

$$\frac{R_c}{R_w} = \frac{\left[\frac{1}{R_w}\right]}{\left[\frac{1}{R_c}\right]} = \frac{\left[\frac{1}{R_{wL}} + \frac{1}{R_{wG}} + \frac{1}{R_{wS}}\right]}{\left[\frac{1}{R_{cL}} + \frac{1}{R_{cG}} + \frac{1}{R_{cS}}\right]} \tag{13}$$

By substituting for the individual phase resistances at the wall and at the bed center in terms of their corresponding heat transfer parameters based on Eqs. 4, 5, and 6, we get,

$$\frac{R_c}{R_w} = \frac{R[h_{wL} + h_{wG} + h_{wS}]}{4[k_{rL} + k_{rG} + k_{rS}]} \tag{14}$$

Similarly, R_c/R_w for the whole bed based on Eq. 2, can also be written as:

$$\frac{R_c}{R_w} = \frac{\left[\frac{R}{4k_{er}}\right]}{\left[\frac{1}{h_w}\right]} = \frac{h_w R}{4k_{er}} = \frac{Bi}{4} \tag{15}$$

From Eqs. 14 and 15, a relation for *Bi*, applicable to two-phase flow, is obtained as:

$$Bi = \frac{R[h_{wL} + h_{wG} + h_{wS}]}{[k_{rL} + k_{rG} + k_{rS}]} \tag{16}$$

So, the final expression for k_{er}, from Eqs. 12 and 16, in terms of the individual phase heat transfer parameters is:

$$\frac{k_{er}}{k_L} = \left[1 + \frac{4(k_{rL} + k_{rG} + k_{rS})}{R(h_{wL} + h_{wG} + h_{wS})}\right]\left[\frac{k_{rS}}{k_L}\left(\frac{Bi_S}{Bi_S + 4}\right) + \frac{k_{rG}}{k_L}\left(\frac{Bi_G}{Bi_G + 4}\right) + \frac{k_{rL}}{k_L}\left(\frac{Bi_L}{Bi_L + 4}\right)\right] \tag{17}$$

Eq. 12, the equivalent form of Eq. 17, readily reduces to the corresponding single-phase thermal resistance model equations derived by Dixon (1985) by putting one of the fluid phase terms to zero (either Bi_L or $Bi_G = 0$), and either $Bi = Bi_F$ for high *Re* case or $Bi = Bi_S$ for low *Re* case, as given by Eqs. 18 and 19 respectively.[8]

High *Re* case:

$$\frac{k_{er}}{k_F} = \frac{k_{rF}}{k_F} + \frac{k_{rS}}{k_F}\left(\frac{Bi_S}{Bi_S + 4}\right)\left(\frac{Bi_F + 4}{Bi_F}\right) \tag{18}$$

Low *Re* case:

$$\frac{k_{\mathrm{er}}}{k_{\mathrm{F}}}=\frac{k_{\mathrm{rS}}}{k_{\mathrm{F}}}+\frac{k_{\mathrm{rF}}}{k_{\mathrm{F}}}\left(\frac{Bi_{\mathrm{F}}}{Bi_{\mathrm{F}}+4}\right)\left(\frac{Bi_{\mathrm{S}}+4}{Bi_{\mathrm{S}}}\right) \tag{19}$$

The predictive equations for single-phase flow based on the thermal resistance models (Eqs. 18 and 19) are identical to those equations obtained by matching the analytical solutions of pseudo-homogeneous model with that of the heterogeneous model by taking $Bi = Bi_{\mathrm{F}}$ for high Reynolds number case (fluid phase dominant) and by taking $Bi = Bi_{\mathrm{S}}$ for low Reynolds number case (solid phase dominant).[15,16] As Eqs. 16 and 17 are also derived based on the thermal resistance networks similar to single-phase flow, they are valid to estimate the heat transfer parameters in two-phase flow through packed beds, irrespective of the flow direction of the fluid phases.

EXPERIMENTAL WORK

Experiments were carried out to obtain the data on radial temperature profile in a trickle-bed reactor (gas–liquid co-current downflow through packed bed reactor). The schematic diagram of the experimental setup is shown in Fig. 3.

The experimental setup mainly consists of a packed bed column of 50 mm diameter, comprising of air–liquid distributor, calming section, jacketed test section and air–liquid separator with other auxiliary parts. Air was drawn from 3.7 kW double piston–double action compressor, of maximum volumetric capacity of 14.98 m3/s at STP and a working pressure of 12 atm. The air drawn from the compressor was saturated with water in a saturator. The saturated air was introduced at the top of the column through a set of pre-calibrated rotameters to the air–liquid distributor at a constant pressure of 4 atm, monitored by a pneumatic pressure regulator. The air flow rates in the rotameter were controlled by needle valves.

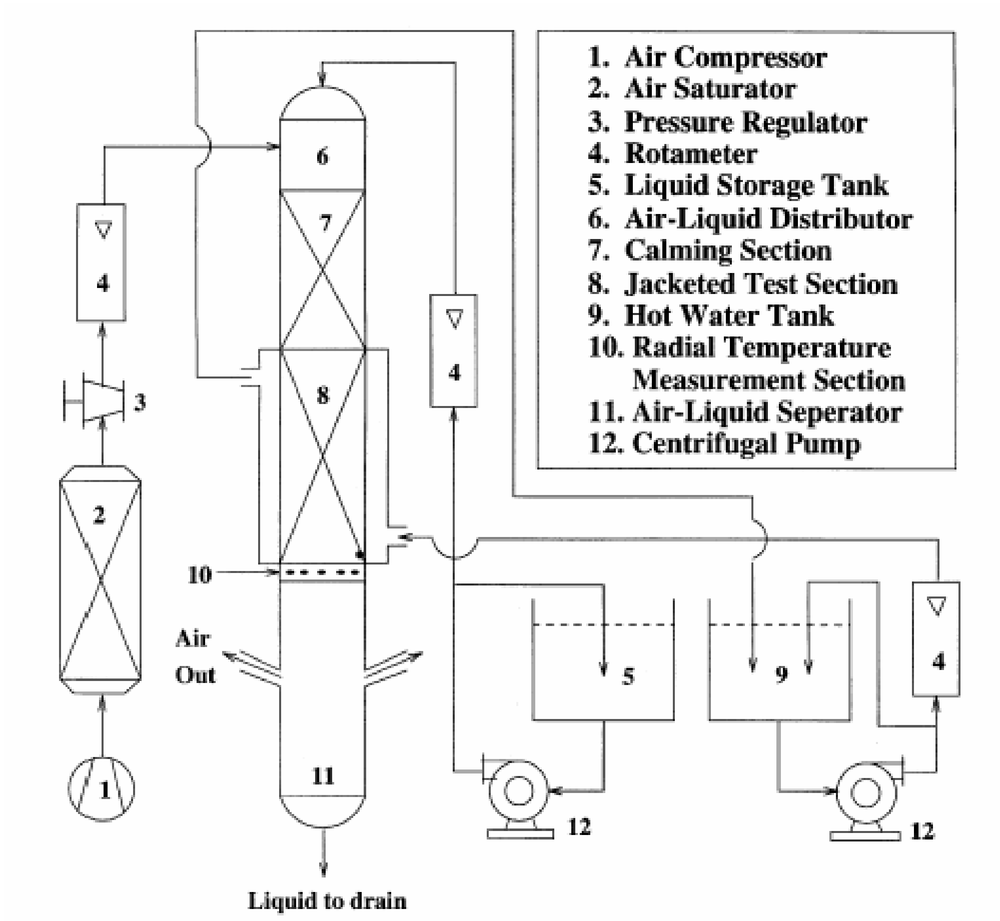

Figure 3. Schematic diagram of Experimental Set-up.

The air was passed through a filter before entering the distributor to remove traces of oil and dust, if any. Water was pumped through a 4.5 kW pump and was metered through a set of pre-calibrated rotameters to the air–liquid distributor at the top of the column. Water flow rates to the column were controlled by means of globe valves. Air and water were uniformly distributed through an air–liquid distributor at the top of the calming section. The air–liquid distributor essentially consists of two sets of openings, 21 copper tubes for distributing liquid and 16 nozzles for distributing air. The tubes and the nozzles were alternately arranged on a triangular pitch over the column cross-section. The number of liquid distribution tubes per unit area was approximately equal over the entire cross-section to ensure equal distribution of liquid. The calming section consisted of a long tube, which ensured a fully developed equilibrium gas–liquid flow before it entered the test section. The jacketed test section was designed for heat transfer studies. It consists of a jacket in order to circulate hot water at 60°C. Hot water was pumped with a 4.5 kW and metered through a set of pre-calibrated rotameters. Below the heat transfer section radial temperature profile measurement section was provided. An air–liquid separator was provided at the bottom of the column to separate the air and the liquid phases coming out of the test section. The radial temperature profile was obtained by measuring the temperatures at the bottom of the test section at three radial positions at r/R = 0.0, 0.4, and 0.8, and at three symmetric angular positions (120° apart) for each radial position. Wall temperature was measured by installing a thermocouple at the inside wall 3 mm above from bottom of the jacketed test section. Thermocouples were also installed at various locations to measure the inlet temperature of test liquid, and inlet and outlet temperatures of hot water.

An INSREF constant temperature bath having an accuracy of 0.01°C and a MINCO platinum resistance thermometer bridge (MINCO RTB8078, Model No. S7929 Pail120C) with an accuracy of 0.025°C as a standard thermometer were used for calibrating all the chormel–alumel thermocouples used in the present study. All these thermocouples were connected to an APTEK multi-channel digital temperature scanner for recording the temperatures. The detailed description of the experimental set-up, and the data collection and reduction procedures are reported elsewhere.[5,13,14] Air and water were fed to the column from the top at the desired flowrates by means of pre-calibrated rotameters. Hot water was circulated through the jacket around the test section at sufficiently high flow rates (25–30 LPM) in order to maintain nearly constant wall temperature, and the minimum and maximum temperature difference between the inlet and the outlet hot water streams were 0.3°C at low flow rates to 2°C at high flow rates respectively of the flowing fluids. In general, it took 20–40 min for attaining the steady state. After steady state was attained, which was confirmed from the constant values of flow rates and temperatures, the flow rates of air and water and the temperatures were recorded. The average of the three angular positions was taken as the temperature at each radial position. This procedure was repeated for a wide range of air (0.01–0.898 kg/m2 s) and water flow rates (3.16– 71.05 kg/m2 s), covering trickle, pulse and dispersed bubble flow regimes. The length of the heat transfer test section used for heat transfer experiments was 0.715 m. The packing employed were 2.59 mm ceramic spheres, 4.05 and 6.75 mm glass spheres and 4.0 and 6.75 mm ceramic raschig rings.

The optimization method in conjunction with the measured radial temperature profile was reported to be an accurate one as compared to the graphical methods to estimate the heat transfer parameters from the two-dimensional pseudo-homogeneous two-parameter model.[7] Therefore, an optimization method employing the IMSL sub-routines BCPOL and ZREAL was used for estimating k_{er} and h_w from the solution of the model equation by minimizing an objective function, as given by Eq. 20.

$$F = \sum_{1}^{M} \left(T_{cal} - T_{exp}\right)^2 \tag{20}$$

where, M is the number of the radial temperature measurement points including the point at the wall. The detailed method of estimating the experimental heat transfer parameters is reported elsewhere.[13,14]

RESULTS AND DISCUSSION

ESTIMATION OF INDIVIDUAL SOLID AND FLUID PHASE PARAMETERS

Eq. 17 involves the individual single-phase heat transfer parameters, and the knowledge of these parameters, viz. k_{rS} and h_{wS} for solid phase; k_{rG} and h_{wG} for gas phase; and k_{rL} and h_{wL} for liquid phase, is required for estimating the two-phase bed heat transfer parameters k_{er} and h_w.

SOLID PHASE PARAMETERS

The parameter k_{rS} is essentially equal to k_{e0}, the stagnant effective radial thermal conductivity of the bed center under no flow conditions. The contribution of k_{rS} and k_{er} for two-phase flow through packed beds is essentially equal to the stagnant effective radial thermal conductivity of the wet bed under no flow conditions, in which the voids are occupied by both gas and liquid. In principle, the Kunii and Smith (1960) equation for k_{e0} or the experimental values of k_{e0} reported in the literature,[17,18,19] valid for single-phase gas flow (when the voids are fully occupied by gas), cannot be used for k_{rS} under two-phase flow conditions.

In view of the good agreement between the experimentally determined k_{e0} values under wet bed conditions and the values of k_{e0} calculated from the theoretical model (Eq. 21) proposed by Babu and Rao (1998), Eq. 21 was used for estimating k_{rS} in this study.[5]

$$\frac{k_{e0}}{k_L} = \varepsilon\left(\frac{k_G}{k_L}\right) + \left[\frac{\beta(1-\varepsilon)}{\phi + \gamma\left(\frac{k_L}{k_S}\right)}\right] \tag{21}$$

The parameter h_{wS} was estimated from Eq. 22, as recommended by Dixon and Cresswell (1979),[15] which was originally proposed by Olbrich (1971).[20]

$$h_{wS} = 2.12\frac{k_{rS}}{d_p} \tag{22}$$

GAS PHASE PARAMETERS

The correlations for estimating the individual gas-phase heat transfer parameters, (i) the gas phase radial thermal conductivity of the bed center, k_{rG}, were reported by Fahien and Smith (1955),[21] Roemer et al. (1962),[22] Gunn and Pryce (1969)[23] and Dixon et al. (1984);[24] and (ii) the gas phase wall-to-bed heat transfer coefficient, h_{wF} were reported by Yagi and Wakao (1959),[25] Kunii and Suzuki (1968),[26] Olbrich and Potter (1972),[27] and Dixon et al. (1984).[24] All these correlations were obtained based on the analogy between heat and mass transfer using mass transfer experiments. In comparison with the other correlations, the Gunn and Pryce (1969) correlation for k_{rG} covers a wide range of experimental conditions for *Re* (0.02-420) and ε (0.26-0.476),[23] and the Yagi and Wakao (1959) correlation for h_{wG} covers a broad range of *Re* (1-2000).[25] Hence their correlations were used for estimating k_{rG} and h_{wG} respectively in the present study.

LIQUID PHASE PARAMETERS

Most of the studies reported for the single-phase heat transfer parameters such as k_{rF} and h_{wF} were based on gas phase as the flowing fluid. Moreover, these parameters were indirectly obtained by analogy from mass transfer measurements.[23,24] The validity of these correlations (reported for k_{rF}) for liquid phase is questionable, because the liquid occupies a part of the voids of the bed and its holdup was not considered in their correlations. Further, no correlations are available for k_{rF} based on heat transfer measurements, be it a gas or a liquid. In fact, the comparison between the experimental values of k_{er} and h_w obtained in the present study with those calculated from the thermal resistance model for two-phase flow (Eq. 17) was not satisfactory, when the liquid phase parameters were estimated using the correlations reported by the above authors. Babu (1993) obtained the individual liquid phase heat transfer parameters, k_{rL} and h_{wL}, from the experiments on k_{er} and h_w for single-phase liquid downflow through packed beds in conjunction with the thermal resistance models applicable to single-phase packed bed heat transfer.[13] The correlations proposed by him for k_{rL} and h_{wL}, given by Eqs. 23 and 24 respectively based on his heat transfer experimental data, were used in the present study.

$$Pe_{rL} = 0.901\left(\mathrm{Re}_L^*\right)^{0.25}\left(\frac{a_v d_p}{\varepsilon}\right)^{0.126}\left(\mathrm{Pr}_L\right)^{1/3} \tag{23}$$

$$Nu_{wL} = \frac{h_{wL} D_e}{k_L} = 0.6\left(\mathrm{Pr}_L\right)^{1/3}\left(\mathrm{Re}_L\right)^{0.47} \tag{24}$$

COMPARISON OF THERMAL RESISTANCE MODEL PREDICTIONS WITH EXPERIMENTAL DATA

From the knowledge of the individual single-phase heat transfer parameters, the effective radial thermal conductivity of the bed center, k_{er}, was calculated from Eq. 17. Then h_w was estimated from the definition of Biot number (Bi), using the known values of Bi from Eq. 16 and k_{er}.

A typical comparison for k_{er} and h_w between the experimental values and the values computed from the thermal resistance model (Eqs. 16 and 17) for 4 mm Raschig rings is shown in Fig. 4, as a function of liquid mass flow rate at a constant gas flow rate of 0.1020 kg/m^{2}s. The agreement is very good between the model predictions and the experimental values with a standard deviation (SD) of 0.06 and 0.08 respectively for k_{er} and h_w respectively.

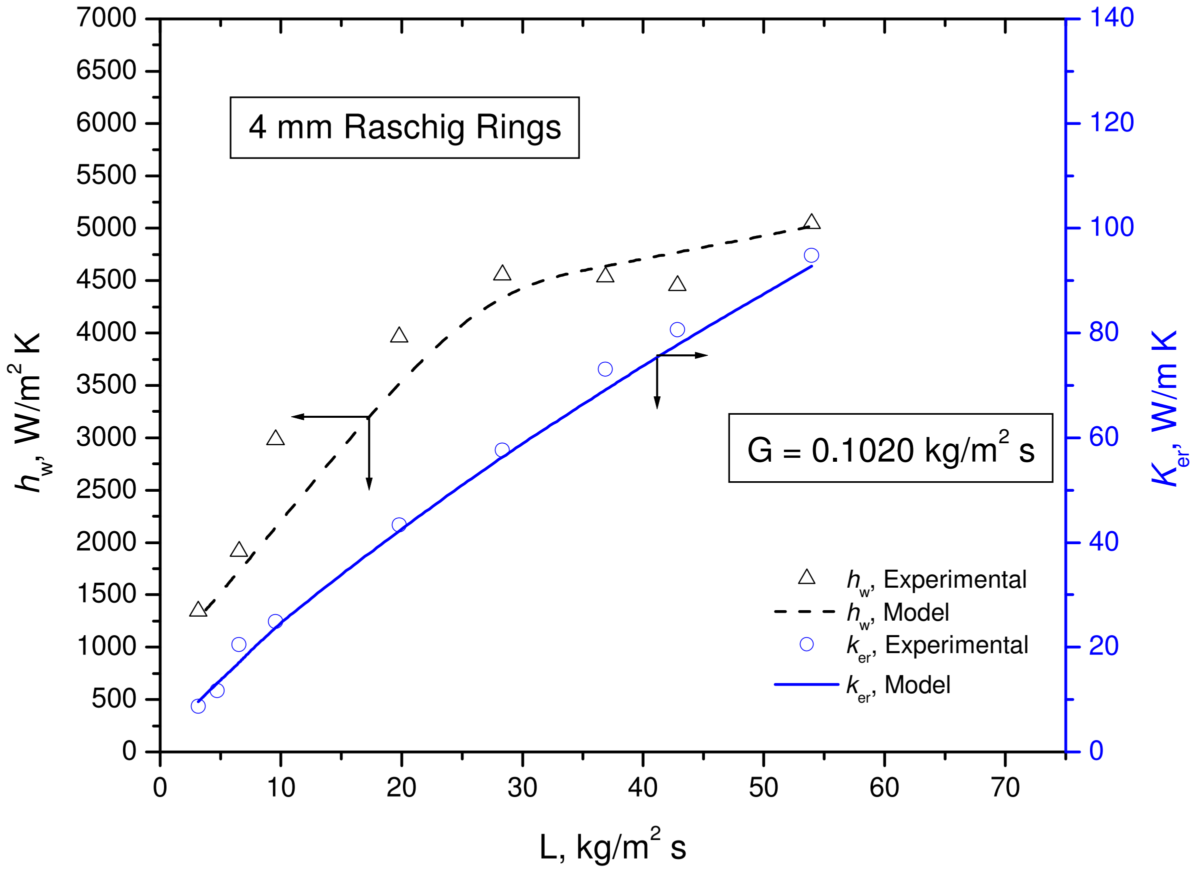

Figure 4. Comparison of the present experimental data on k_{er} and h_w in two-phase flow with the values predicted from the thermal resistance model as a function of liquid mass velocity.

The comparison of the calculated values from the thermal resistance model with the entire present experimental data and the experimental data reported by Specchia and Baldi (1979),[4] Matsuura et al. (1979b),[28] and Crine (1982)[29] on k_{er} is shown in Fig. 5 and Fig. 6 respectively. Most of the present experimental data lies within ± 25% error limits with a SD of 0.14, while most of the literature data in Fig. 6 lies within the error limits of -15 to +25% with a SD of 0.12.

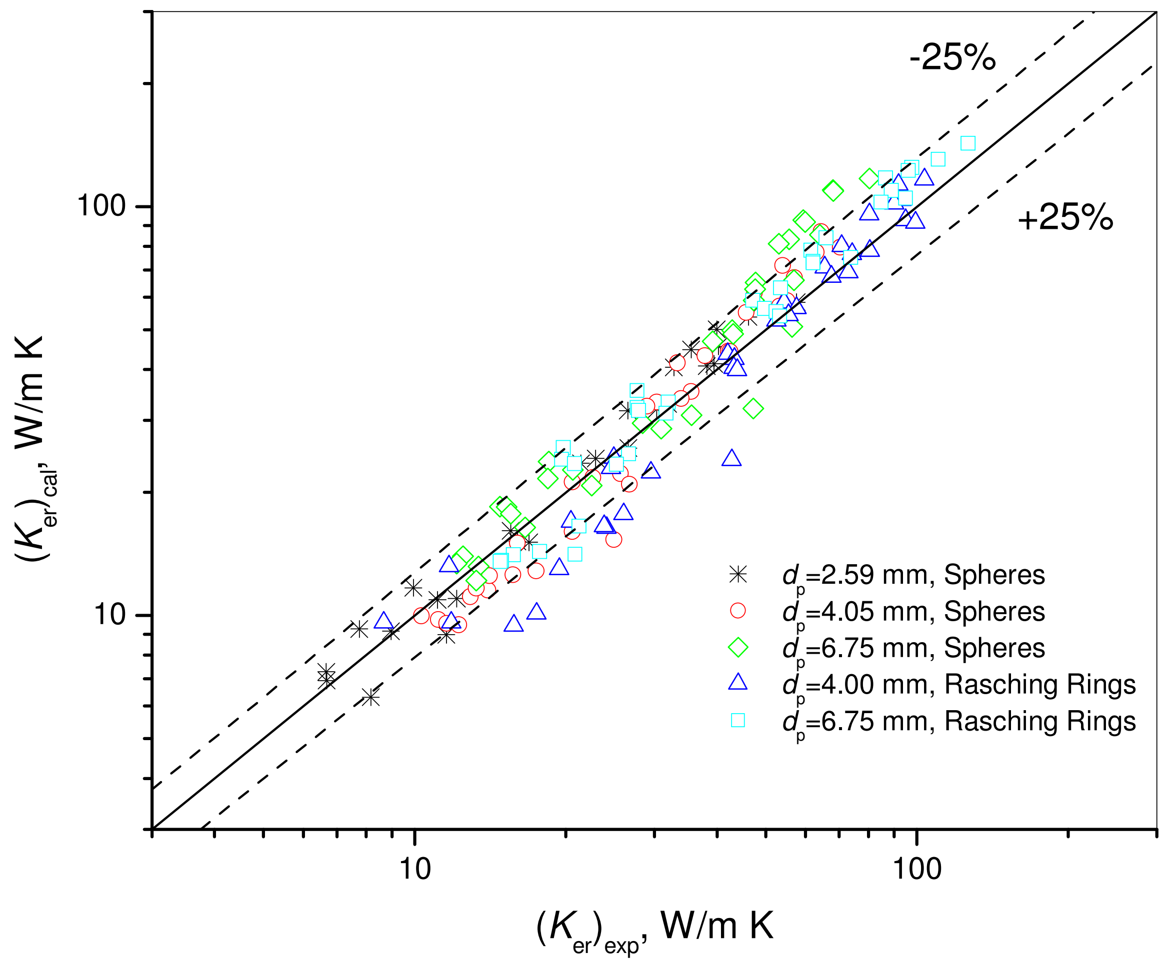

Figure 5. Comparison of the present experimental data on k_{er} with the values calculated from the thermal resistance model for two-phase flow.

The present experimental data and that reported in the literature on h_w are also compared with the model predictions in Fig. 7 and Fig. 8 respectively. The model predicts comparable h_w values within the error limits of -15 to +35% for present experimental data with a SD of 0.17, whereas it predicts within -25 to +35% error limits for the literature data[4,12] with a SD of 0.19.

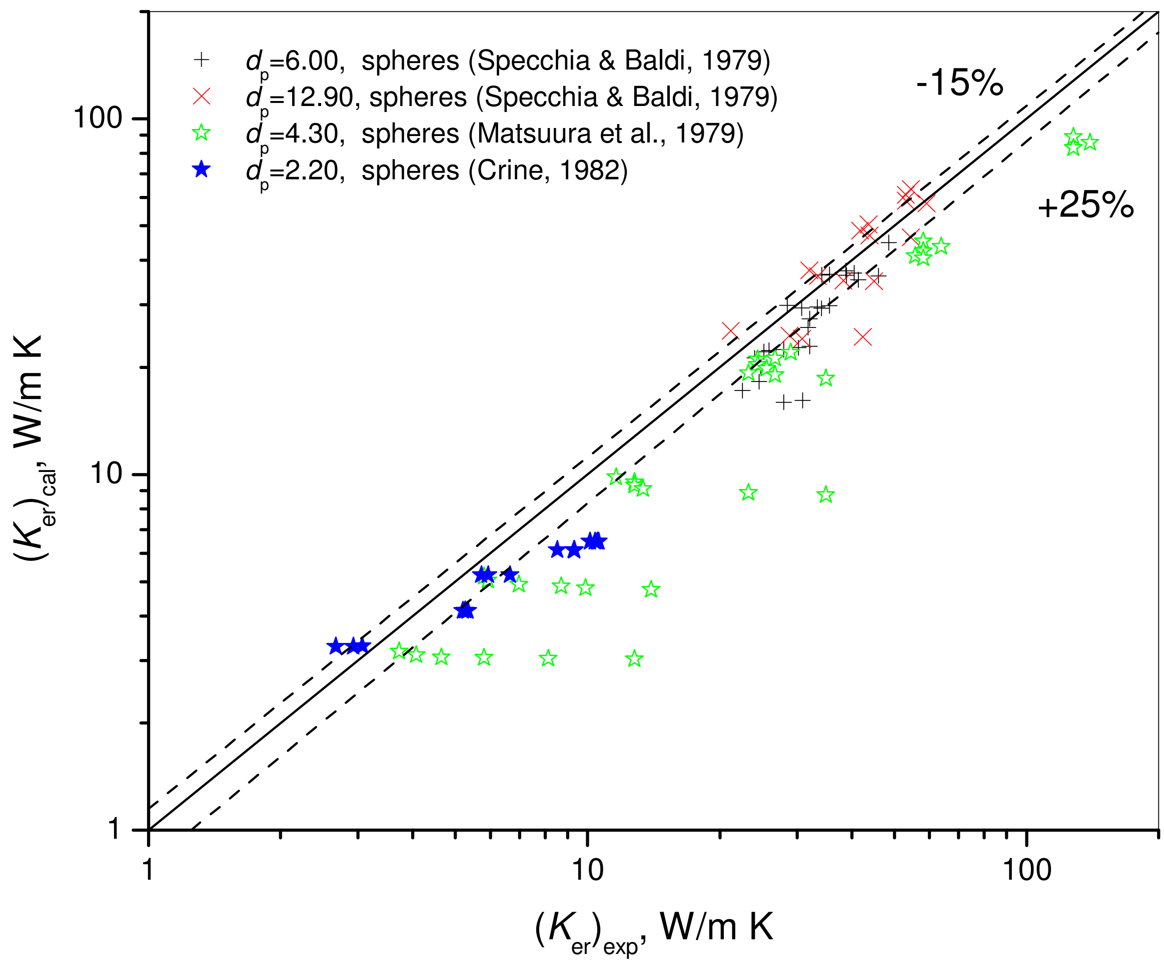

Figure 6. Comparison of the literature experimental data on k_{er} with the values calculated from the thermal resistance model for two-phase flow.

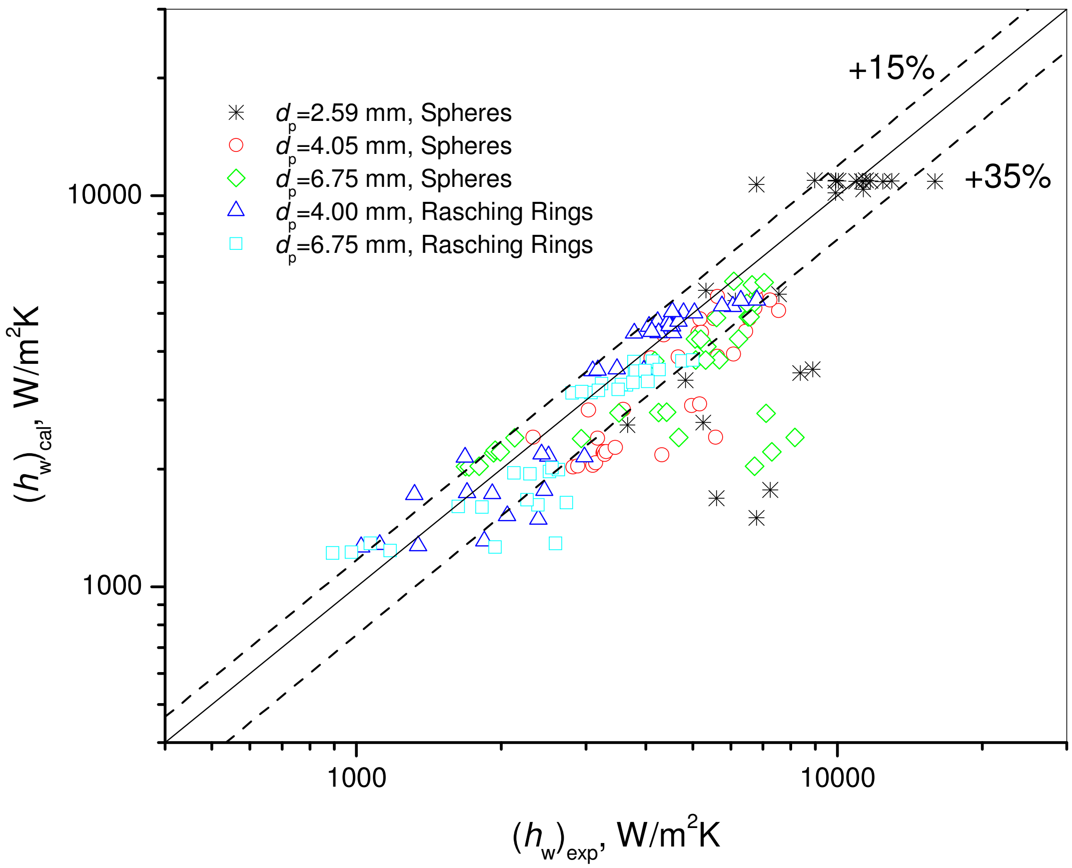

Figure 7. Comparison of the present experimental data on h_w with the values calculated from the thermal resistance model for two-phase flow.

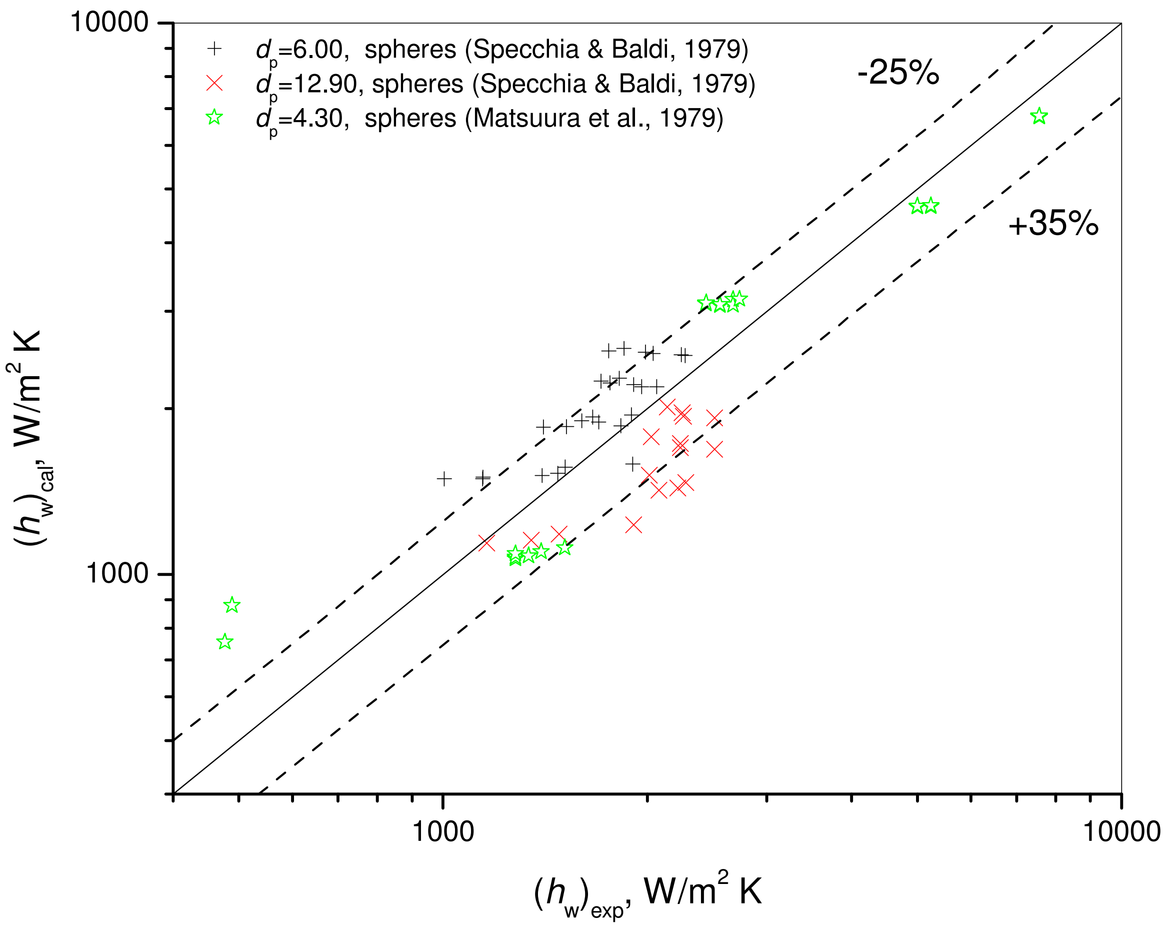

Figure 8. Comparison of the literature experimental data on h_w with the values calculated from the thermal resistance model for two-phase flow.

CONCLUSIONS

1. The predictive equations for k_{er} and h_w for two-phase flow through packed beds were derived using thermal resistance networks.
2. The reported correlations for estimating the individual fluid phase heat transfer parameters, which were based on the mass transfer analogy and gas as the flowing fluid were found to be inadequate for estimating the liquid-phase heat transfer parameters.
3. The comparison of the experimental heat transfer parameters, k_{er} and h_w, of the present study and those reported in the literature, with the values calculated from the proposed thermal resistance model for two-phase flow was found to be satisfactory.

NOTATION

ROMAN SYMBOLS

a_v specific surface area of the particle, $6/D_e$, m^{-1}

Bi Biot number, $h_w R/ k_{er}$, (-)

D_e equivalent particle diameter, defined as the diameter of a sphere having the same surface area to volume ratio as that of the particle, m

d_p nominal particle diameter, m

h_m mean or overall heat transfer coefficient, W/m^2K

h_w wall-to-bed heat transfer coefficient, W/ m^2K

h_{w0} wall-to-bed heat transfer coefficient of stagnant bed under no flow conditions, W/ m^2K

h_{wF} wall-to-bed heat transfer coefficient of fluid, W/ m^2K

h_{wS} wall-to-bed heat transfer coefficient of solid, W/ m^2K

k thermal conductivity, W/mK

$\bar{k}_r$ average effective radial thermal conductivity, W/mK

k_{e0} effective radial thermal conductivity of stagnant bed under no flow conditions, W/mK

k_{er} effective radial thermal conductivity of the bed, W/mK

$\bar{k}_{er}$ average effective radial thermal conductivity of the bed, W/mK

k_{rF} effective radial thermal conductivity of fluid, W/mK

k_{rG} effective radial thermal conductivity of gas, W/mK

k_{rL} effective radial thermal conductivity of liquid, W/mK

k_{rS} effective radial thermal conductivity of solid, W/mK

Nu Nusselt number, $h_w D_e/ k$, (-)

n number of data points used in estimating the standard deviation

Pe Peclet number, $U\rho C_p d_p/k$, (-)

Pe_{rF} radial Peclet number of fluid, $LC_p d_p/k_{rF}$, (-)

Pr Prandtl number, $C_p\mu/k$, (-)

R radius of the column (when unsubscripted), m

R thermal resistance per unit wall area (when subscripted), m^2K/W

Re Reynolds number, $D_e U\rho/\mu$, (-)

$Re*$ Modified Reynolds number, $Re/(\varepsilon\beta_t)$, (-)

r radial position, m

U superficial velocity of the fluid, m/s

GREEK SYMBOLS

β liquid holdup, (-)

ε bed voidage or porosity, (-)

Φ dimensionless thickness of a slab of stationary liquid at the contact points between the particles in the bed, (-)

γ dimensionless thickness of a slab of solid material of the packing in the bed, (-)

μ viscosity, kg/m-s

ρ density, kg/m^3

SUBSCRIPTS

c center

F fluid

G gas

L liquid

p particle

S solid

t total

w wall

ABBREVIATIONS

cal calculated

exp experimental

SD standard deviation, $\sqrt{\dfrac{\sum_{1}^{n}\left[\dfrac{Y_{\exp}-Y_{\mathrm{cal}}}{Y_{\exp}}\right]^{2}}{n-1}}$

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