HEAT TRANSFER IN A MEMBRANE ASSISTED FLUIDISED BED WITH IMMERSED HORIZONTAL TUBES

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ABSTRACT

The effect of gas permeation through horizontally immersed membrane tubes on the heat transfer characteristics in a membrane assisted fluidised bed was investigated experimentally. Local time-averaged heat transfer coefficients from copper tubes arranged in a staggered formation with the membrane tubes to the fluidised bed were measured in a square bed (0.15 m x 0.15 m x 0.95 m) containing glass particles (75-110 µm) fluidised with air distributed via a porous plate, where the ratio of gas fed or removed through the membrane bundles and the porous plate distributor was varied. The experimental results revealed that high gas permeation rates through the membranes strongly decreased the heat transfer coefficient at high superficial gas velocities for tubes at the top of the tube bundle, which was attributed to the reduced mobility and heat capacity (higher dilution) of the emulsion phase. However, the effect of gas permeation was much less pronounced for tubes lower in the tube bundle because of the lower local dilution of the emulsion phase.

INTRODUCTION

Fluidised beds employing fine powders are finding increased application in the chemical and petrochemical industry because of their excellent mass and heat transfer characteristics. However, in fluidized bed chemical reactors axial gas back-mixing can strongly decrease the conversion and product selectivities. By insertion of membranes in fluidized beds large improvements in conversion and selectivity can be achieved, firstly by optimizing axial concentration profiles via distributive feeding of one of the reactants or selective withdrawal of one of the products, and secondly, by decreasing the effective axial dispersion via compartmentalization of the fluidized bed. Moreover, insertion of membrane bundles in a suitable configuration impedes bubble growth, thereby reducing reactant by-pass via rapidly rising large bubbles. Often cooling or heating tubes are also submerged in the fluidised bed to withdraw or add thermal energy. The effective heat transfer coefficient between the surface of these tubes and the fluidised bed is an important parameter in the design of these fluidised beds. The integrated gas addition or removal via membranes inside the fluidised bed strongly influences the bed hydrodynamics and thus the tube-to-bed heat transfer. In this work the influence of the presence of membrane and heat transfer tube bundles and the effect of gas addition and removal via the membrane tubes on the spatial distribution of the tube-to-bed heat transfer coefficient was studied experimentally.
EXPERIMENTAL SET UP

To measure the spatial distribution of the heat transfer coefficient, a square fluidised bed (0.15 m x 0.15 m x 0.95 m) was constructed out of lexan and filled with massive glass beads (75-110 µm, 2550 kg/m³) to a packed bed height of about 0.30 m. The bed was equipped with 18 horizontal copper heat transfer tubes (2 mm ID and 3 mm OD) and 40 horizontal ceramic membrane tubes (1.5 mm ID and 2.5 mm OD with a pore size of 0.15 µm), through which gas could be added or withdrawn, arranged in a staggered arrangement with an equilateral pitch of 0.02 m. A schematic diagram of the experimental set-up is shown in Fig.1. Uniform fluidisation was achieved with a porous plate distributor with a pore size of 10 µm. Fluidisation was performed with humidified air (50-60 % humidity) at ambient conditions to avoid static electricity problems. It was found that static charging of the particles strongly decreased the measured heat transfer coefficients, especially in experiments with tube bundles. Furthermore, it was verified experimentally that the bed temperature was uniform throughout the bed.

Fig. 1. Schematic of the experimental set-up

EXPERIMENTAL TECHNIQUE

Various experimental methods have been reported in the literature to measure the heat transfer coefficient between a surface and the fluidised bed, some use an electrically heated metallic film sensor (e.g. Tout and Clift (1) and Fitzgerald et al. (2)), other use thermocouples (e.g. George (3), Olsson and Almsted (4), McKain et al. (5), Karamavme et al. (6) and Kahn and Turton (7)). In this work the tube-to-bed heat transfer coefficient was determined from the difference in the entrance and the exit mixing-cup temperatures of the heat transfer tubes, which were fed with preheated water (4.6 cm/s) to about 50 °C, and the bed temperature using T-type thermocouples. An advantage of this technique is that the thermocouples can be switched easily from one tube to another to determine the axial and lateral variation in the heat transfer coefficient.
A thermal energy balance over a heat transfer tube submerged in the fluidised bed reads:

\[
\phi_m C_p \frac{dT_{\text{water}}}{dz} = \pi d_i h_{\text{total}} (T_{\text{water}} - T_{\text{bed}})
\]

where the overall heat transfer coefficient, \(h_{\text{total}}\), is given by:

\[
\frac{1}{h_{\text{total}}} = \frac{1}{h_{\text{water}}} + \frac{d_i \ln\left(\frac{d_i}{d_o}\right)}{2 \lambda_{\text{copper}}} + \frac{1}{d_o h_{\text{bed}}}
\]

Assuming constant physical properties and a constant fluidised bed temperature, and taking \(Nu = h_{\text{water}}d_i/\lambda_{\text{water}} = 3.66\) for the heat transfer resistance in the laminar flow inside the copper tube, the measured axial temperature profile inside the heating tube could be described well. With this technique the tube-to-bed heat transfer coefficient could be determined within an experimental error of maximum ca. 10%.

**EXPERIMENTAL RESULTS AND DISCUSSION**

Firstly, experimental results on the tube-to-bed heat transfer coefficient for a single tube are presented and compared with reported experimental values in the literature. Subsequently, the results for tube banks without permeation through the membranes will be reported. Finally, the effect of permeation through the membranes will be discussed.

**Heat Transfer From A Single Tube**

The heat transfer coefficient between the surface of a single tube submerged in a fluidised bed was measured at different positions in the fluidised bed in order to compare the observed heat transfer coefficients with reported literature values and as a reference for the experiments employing tube-banks. The experimentally determined tube-to-bed heat transfer coefficient increased with increasing superficial gas velocities and reached a maximum at about 8 \(u_{mf}\). The maximum tube-to-bed heat transfer coefficient (\(h_{\text{max}}\)) increased as a function of height above the distributor (position 16 (see Fig.1): 830 W/m²K; position 2: 970 W/m²K), which is attributed to increased solids mobility higher in the bed due to bubble coalescence. In the lateral direction no significant changes in the heat transfer coefficient were observed even at high superficial gas velocities. Wall effects were not measured since the measurement closest to the wall was 1.7 cm. The measured tube-to-bed heat transfer coefficients compare well with experimental values reported in the literature, which were measured with heat transfer probes for similar systems under comparable fluidization conditions (see Table 1).

<table>
<thead>
<tr>
<th>Material</th>
<th>(d_p) ((\mu m))</th>
<th>(h_{\text{max}}) (W/m²K)</th>
<th>(h_{\text{avg}}) (W/m²K)</th>
<th>Reference</th>
</tr>
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<tr>
<td>Glass</td>
<td>76</td>
<td>-</td>
<td>766</td>
<td>Sharma, 1997 (8)</td>
</tr>
<tr>
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<td>76</td>
<td>-</td>
<td>825</td>
<td>Sharma and Turton, 1998 (9)</td>
</tr>
<tr>
<td>Glass</td>
<td>100</td>
<td>850</td>
<td>-</td>
<td>Sharma and Turton, 1998 (9)</td>
</tr>
</tbody>
</table>
Heat Transfer With A Tube Bank Without Membrane Permeation

The tube-to-bed heat transfer coefficients were measured in the fluidised bed for all the 18 tubes placed in the bed at different superficial gas velocities without permeation through the membrane tubes. As shown in Fig. 2, the measured heat transfer coefficient increases with the superficial gas velocity and levels off at higher gas velocities to a maximum. Increasing the superficial gas velocity increases the mobility of the emulsion phase, which decreases the average residence time of an emulsion phase ‘packet’ at the tube surface, thereby increasing the heat transfer coefficient. However, at higher gas velocities larger bubbles are formed having a lower volumetric heat capacity, causing the heat transfer coefficient to level off and eventually decrease at very high gas velocities (see e.g. Kunii and Levenspiel (10)). Furthermore, the heat transfer coefficient strongly decreases as a function of the distance from the distributor, caused by the bubble growth.

When comparing the maximum heat transfer coefficient determined in the fluidised bed with a tube bank with the results obtained with a single tube, the tube-to-bed heat transfer coefficient decreased by almost 200 W/m²K (see Fig. 2.b) due to the reduced mobility of the emulsion phase caused by the additional internals, which obstruct the macro-scale movement of the emulsion phase. Moreover, the internals cause bubble breakage, which reduces the bubble rise velocity, resulting in decreased emulsion movement at the heat transfer surfaces.

Fig.2. Tube-to-bed heat transfer coefficient as a function of the superficial gas velocity
a) For different heights above the distributor (Tube number refers to position indicated in Fig.1); b) for an experiment with a single tube and a tube bank, measured at position 2.

In Fig. 3 the spatial distribution of the heat transfer coefficient for two different superficial gas velocities is shown. Increasing the superficial gas velocity from 6 $u_{mf}$ to 10 $u_{mf}$ increases the tube-to-bed heat transfer coefficient, but does not change its spatial distribution. The highest heat transfer coefficients were observed just above the distributor in the center of the bed. A slight lateral asymmetry in the spatial distribution is caused by the asymmetric configuration of the heat transfer tubes.
Fig. 3. Spatial distribution of the tube-to-bed heat transfer coefficient for two different superficial gas velocities: (a) $6u_{mf}$ and (b) $10u_{mf}$.

Heat Transfer With A Tube Bank With Membrane Permeation

To study the effect of gas permeation via membranes on the tube-to-bed heat transfer coefficient, experiments were carried out by adding and removing part of the fluidising gas via the membranes at different superficial gas velocities. Up to 40% of the total gas flow could be added via the membranes, whereas only 10% of the total gas flow could be removed due to pump limitations. In the experiments where gas was added through the membranes, the total gas feed was kept constant, which implies that experiments with higher permeation rates through the membranes were carried out at a lower gas flow through the distributor.

In Fig. 4 the spatial distribution of the tube-to-bed heat transfer coefficient is given at a superficial gas velocity of $6u_{mf}$ for different membrane permeations. The Figure clearly shows that with increasing gas permeations through the membrane the measured heat transfer coefficient at the bottom of the bed decreases and that the heat transfer coefficient decreases much more pronounced as a function of the axial position in the bed. The lower heat transfer coefficient at the bottom of the bed at higher gas permeations was caused firstly by the lower gas feed through the distributor and secondly by the suppressed macroscopic circulation pattern due to the reduced downflow at the walls and the reduced bubble growth in the centre of the bed. Furthermore, the heat transfer coefficient decreases strongly as a function of the height above the distributor and even much more pronounced than observed for the experiment without permeation, where the decrease in the heat transfer coefficient was caused by the bubble growth. The additional decrease in the heat transfer coefficient as a function of the axial coordinate is attributed to the dilution of the emulsion phase with the gas fed via the membranes resulting in a decreased heat capacity of the emulsion phase.

Fig. 5 depicts the effect of the superficial gas velocity on the tube-to-bed heat transfer coefficient at different permeations and at two different tube locations. For the tube located at the top of the bed, the effect of membrane permeation on the tube-to-bed heat transfer coefficient was negligible at a low fluidization velocity of $2u_{mf}$ but very strong at higher gas velocities (see (Fig. 5.a). At a low superficial gas velocity the emulsion packet renewal rate at the tube surface was very low due to the absence of a large macroscopic circulation pattern caused by the absence of larger
bubbles. Hence, the tube-to-bed heat transfer coefficient will mainly depend on the local superficial gas velocity. Remarkably, at high gas permeations through the membrane an increase in the total gas flow does not increase the heat transfer coefficient for a tube at the top of the bundle. The increased macroscopic emulsion circulation at higher gas velocities (because of the larger bubbles) is more than counterbalanced by the increased dilution of the emulsion phase.

For a tube in the center of the bed the effect of permeation through the membranes is very pronounced at a low fluidization velocity of 2 $u_{mf}$ (see Fig. 5.b) because of the reduced local gas velocity at higher permeation rates. However, at high superficial gas velocities only a small decrease in the heat transfer coefficient at higher membrane permeations was observed, because the smaller local dilution of the emulsion phase, since the tube in the center of the tube bundle experiences only part of the total gas fed via membranes.

Fig. 4. Heat transfer coefficients of the bed at various positions in the bed at 6 $u_{mf}$
(a) -10 % permeation; b) no permeation; c) 20% permeation; d) 40% permeation.
Fig. 5. Tube-to-bed heat transfer coefficient as a function of the superficial gas velocity at
a) Tube position 2 (top of the tube bundle); b) Tube position 11 (center of the tube bundle).

CONCLUSIONS

The effect of gas permeation through horizontally submerged membrane tubes in a fluidised bed on the tube-to-bed heat transfer coefficient was investigated experimentally by measuring the bed temperature and the inlet and outlet mixing-cup temperatures of water flowing through heat transfer tubes. The presence of the membrane and heat transfer tubes decreases the heat transfer coefficient due to the reduced mobility of the emulsion phase, caused by the additional internals obstructing the macro-scale movement of the emulsion phase, and by bubble breakage decreasing the bubble rise velocity. Without gas permeation through the membranes, the heat transfer coefficients increase with increasing superficial gas velocity and level off at high gas velocities to a maximum, where the increased emulsion phase mobility is counterbalanced by the larger bubble growth. The heat transfer coefficient decreased as a function of the distance from the distributor, which was attributed to bubble growth. At high gas permeation rates through the membranes the decrease in the heat transfer coefficient at the top of the tube bundle was even stronger at high superficial gas velocities, caused by the additional dilution of the emulsion phase. However, lower in the tube bundle the decrease in the heat transfer coefficient was much less pronounced due to the lower dilution of the emulsion phase, since these tubes experienced only part of the total gas fed via membranes. By controlled dozing of one of the reactants in a membrane assisted fluidised bed the product selectivity and/or operational safety can be enhanced, but care must be taken to include the effect of gas addition through the membranes on the required heat transfer surface area.

ACKNOWLEDGEMENT

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NOTATIONS

\[ C_p \] heat capacity (J/kg K)
\( d_i \) inside diameter of the copper tube (m)
\( d_o \) outside diameter of the copper tube (m)
\( h_{\text{bed}} \) bed heat transfer coefficient (W/m\(^2\) K)
\( h_{\text{total}} \) overall heat transfer coefficient (W/m\(^2\) K)
\( h_{\text{tube}} \) tube side heat transfer coefficient (W/m\(^2\) K)
\( \lambda_{\text{copper}} \) Thermal conductivity of copper (W/m K)
\( \lambda_{\text{water}} \) Thermal conductivity of water (W/m K)
\( L \) Tube length (m)
\( T \) temperature (°C)
\( u \) Superficial gas velocity (m/s)
\( u_{mf} \) minimum fluidisation velocity (m/s)
\( Z \) bed height (m)
\( Nu \) Nusselt number
\( Gz \) Graetz number
\( \phi_m \) mass flow rate of water (kg/s)
\( \rho \) density of water (kg/m\(^3\))

REFERENCES