INFLUENCE OF HYDRODYNAMICS ON HEAT TRANSFER IN FLUIDIZED BEDS

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ABSTRACT

Simultaneous measurements of local bubble behaviour and time
averaged local heat transfer at a vertical single tube and a ver­
tical tube bundle were carried out in fluidized beds of different
diameters. The evaluation of a theoretical heat transfer equation
describing time-averaged heat transfer as a function of solid
particle contact time fits the measurements.

INTRODUCTION

Several theories of the heat transfer process have been
proposed for example Ernst (1959), Mickley et al. (1961), Botterill
and Williams (1963) and Schlünder (1971). In many of the theo­
retical models, the heat transfer is postulated to depend on the
contact time of the particles at the heat transfer surfaces.
To date, useful application of these theories has been prevented
by lack of information regarding particle contact times in flu­
idized beds.

EXPERIMENTAL EQUIPMENT

Three fluidized beds of 0.19, 0.4 and 1.0 m in diameter were
used. An electric heating section with a length of 0.15 m was
mounted in one of the tubes. Three thermocouples were positioned
at different locations to measure the temperature of the heat
transfer surface. The bed temperature was measured with separate
thermocouples and was found to be uniform throughout the bed.
Values of α were calculated according to Eq. (1):
\[
\alpha = \frac{q}{\left[ S \left( T_w - T_e \right) \right]}
\]  

(See list of symbols at the end of the paper).

The bubble behaviour was measured using a capacitance probe developed by Werther and Molerus (1973), which was mounted in the direct vicinity of the heat transfer surface. The following parameters of the local bubble behaviour were determined: the local bubble frequency \( f_B \), the local mean bubble rise velocity \( U_B \), the local mean bubble volume fraction \( \varepsilon_B \).

**EXPERIMENTAL PROGRAM**

Five types of solids were fluidized with air. Dimensions and properties for the five types of test particles are given in Table 1. The ratio of static bed height \( H \) to column diameter \( D \) was 1.5 in all runs. In the fluidized beds of 0.4 and 1.0 m diameter, two types of distributor, a porous plate and a nozzle distributor were used. The 1.0 m diameter fluidized bed could be equipped with a vertical tube bundle consisting of 91 tubes with OD of 0.042 m.

**THEORY**

The time-averaged heat transfer coefficient for heat transfer between a heating surface of constant temperature and a solid body at distance \( \delta_g \) as function of contact time is, according to Ernst (1959):

\[
\alpha = \frac{c_p \rho_p t^{-1}}{(2/\sqrt{\pi}) \sqrt{a_p} + h^{-1} \left( \exp \left( h^2 a_p t \phi_c \right) \frac{\lambda_c}{h a_p} \right) - h^{-1}}
\]

where \( h = \lambda_g / (\delta g_p) \), \( \phi_c = 1 - \phi \), \( a_p = \lambda_p / (\rho_p c_p) \).

<table>
<thead>
<tr>
<th>Solid particles</th>
<th>Surface mean diameter (( \mu ))</th>
<th>Particle density ( \rho_p ) (kg/m(^3))</th>
<th>Minimum fluidizing velocity ( U_{mf} ) (m/s)</th>
<th>Porosity ( \varepsilon_{mf} )</th>
<th>Thermal conductivity ( \lambda_E ) (W/m(^\circ)K)</th>
<th>Specific heat ( C_p ) (Ws/kg(^\circ)K)</th>
</tr>
</thead>
<tbody>
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