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MASS TRANSFER OF OXYGEN TO A BURNING PARTICLE IN A FLUIDIZED BED

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SUMMARY: A theory is outlined for predicting the mass transfer of a gas through a fluidized bed of small non-reacting particles to a large reacting particle. It is proposed that mass transfer of gas takes place by two mechanisms: (i) by transfer of gas in clusters of smaller particles approaching the large particle and (ii) by gas convection. The theory developed enables the prediction of the mass transfer coefficient, expressed as the Sherwood number, in terms of the fluidized-bed operating conditions. The expression derived is shown to be in excellent agreement with Sherwood numbers obtained from experiments in which single large petroleum coke particles were burned in air in fluidized beds of sand.

1. NOTATION

B	constant, equation (1)
C_A	concentration of A
C_{Ao}	C_A at surface
$C_{A\infty}$	C_A far from surface
D	diffusivity of A
D_a	diffusivity of A in air
d	active particle diameter
d_s	inert particle diameter
F_A	mass flux of A
g	acceleration due to gravity
k	constant, equation (9)
N_{Ar}	Archimedes number
N_{Re}	particle Reynolds number
N_{Sc}	Schmidt number
N_{Sh}	Sherwood number
r	radial distance
U	superficial fluidizing velocity
U_{mf}	minimum fluidizing velocity
u_b	absolute bubble rise velocity
β	mass transfer coefficient
ε	voidage at U
ε_{mf}	voidage at U_{mf}
μ	viscosity of medium
ρ	density of medium
ρ_p	density of bed particle
θ	time
θ_m	contact time

2. INTRODUCTION

Prediction of the combustion rate of coal particles in a fluidized bed requires a knowledge of the effect of the non-combustible bed particles on the diffusion rate of oxygen to the burning particle. This effect is characterized by the mass transfer coefficient, β, usually expressed in a dimensionless form known as the Sherwood number, $N_{Sh} = \beta d/D$. For diffusion to a single spherical particle in an extensive fluid, N_{Sh} is often expressed (Rowe et al., 1965) as a function of the flow rate through the Reynolds number ($N_{Re} = Ud\rho/\mu$) and the fluid properties by the Schmidt number ($N_{Sc} = \mu/\rho D$) in the form

$$N_{Sh} = 2 + B\, N_{Re}^{1/2}\, N_{Sc}^{1/3} \qquad (1)$$

where B, an empirical constant, generally taken as equal to 0.69, has been shown to be a slowly varying function of Reynolds number, the level of turbulence and the sphere diameter (Galloway and Sage, 1964).

For large particles (~ 10 mm) burning in a fluidized bed of smaller non-combustible particles (~ 1 mm), N_{Re} values are of the order of 30 which suggests $N_{Sh} \approx 6$, if equation (1) can be applied to the particulate system found in a fluidized bed. No proven relationship is available for predicting N_{Sh} for a burning particle in a fluidized bed of non-combustible particles. This paper outlines a new approach to predicting N_{Sh} for a large burning particle and compares it to a semi-empirical equation proposed recently by the authors (La Nauze and Jung, 1982).

3. EXISTING RELATIONSHIPS FOR N_{Sh} FOR FLUIDIZED BEDS

La Nauze and Jung (1982) measured the combustion rate of single petroleum coke spheres in air-fluidized beds of sand. They proposed that the semi-empirical equation developed by Frössling (1938) and Rowe et al. (1965) for a single isolated particle in the absence of the fluidized bed could be modified by the addition of the bed voidage, ε, to give:

$$N_{Sh} = 2\varepsilon + 0.69 \left(\frac{N_{Re}}{\varepsilon}\right)^{1/2} N_{Sc}^{1/3} \qquad (2)$$

Fluid properties associated with the boundary layer are evaluated at the mean temperature between the bed and the particle whilst the bulk properties such as velocity are evaluated at the bed temperature throughout this paper.

Data presented by La Nauze and Jung (1982) clearly demonstrated that N_{Sh} varied significantly with the changing diameter during burn-off. Figure 1 compares further data with equation (2) and with relationships previously proposed by Chakraborty and Howard (1981) and Pillai (1981), respectively:

$$N_{Sh} = 2\varepsilon + 0.69\, N_{Re}^{1/2}\, N_{Sc}^{1/3} \qquad (3)$$

$$N_{Sh} = 2\varepsilon + 0.69\varepsilon\, N_{Re}^{1/2}\, N_{Sc}^{1/3} \qquad (4)$$

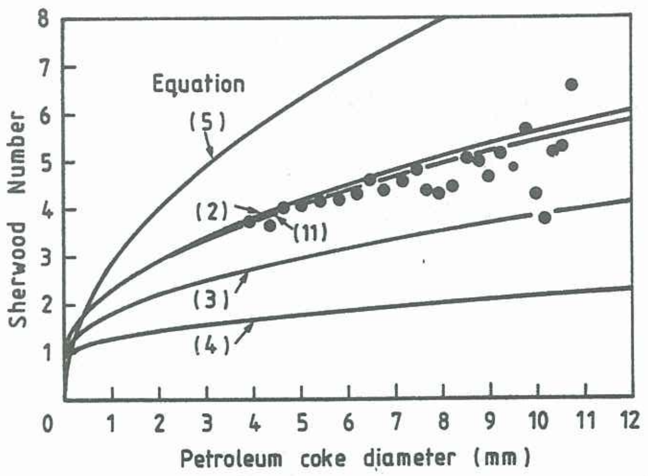

Fig. 1 Comparison between relationships for N_{Sh} and experimental values. Experimental conditions: single petroleum coke particle burning in air-fluidized bed of 0.78 mm sand at 1173 K, U = 0.39 m/s

Tamarin (1981) applied Prandtl boundary layer theory to determine N_{Sh}. The average tangential stress at the particle surface was related to the particle properties to determine the effective Reynolds number to give:

$$N_{Sh} = 0.248\,(N_{Sc}\,N_{Ar})^{1/3}\,(\frac{d}{d_s})^{1/2} \qquad (5)$$

where N_{Ar} is the Archimedes number $(g\,d_s^3\,\rho(\rho_p-\rho)/\mu^2)$ and d_s the average non-combustible particle size.

Comparison between these semi-empirical relationships and the data given in Figure 1 suggests that equation (2) provides the best estimate of the variation of N_{Sh} with d.

4. PROPOSED THEORY FOR N_{Sh} IN FLUIDIZED BEDS

The assumption of the above relationships is that the inert particles merely "shield" the active particle from the gas. It is well established that the solid particles play an important role in heat transfer to a surface immersed in a fluidized bed. Transient heat transfer is effected by "clusters" or "packets" of particles which are periodically displaced from the surface by gas bubbles passing in the vicinity (Gelperin and Einstein, 1971). La Nauze et al. (1983) proposed that these packets also transfer fresh gas to the surface and that the total mass transfer of gas to an active particle is made of:

(i) the mass transfer of gas percolating through the bed at minimum fluidizing conditions (the gas convective component).
(ii) the mass transfer of gas in clusters of particles (packets) that circulate within the bed under the influence of the bubbles (the particle convective component).

An idealised picture of this process is shown in Figure 2 which illustrates the packets forming around a bubble and breaking away from the wake. It is proposed that this concept of mass diffusion is best treated by non-steady state diffusion at the gas-solid interface.

This approach, called the penetration theory (Higbie, 1935), deals with the non-stationary nature of the medium by assuming that fresh gas presents itself from time to time to the gas/solid interface.

The solution for non-steady state diffusion to a sphere of radius d/2 is detailed by La Nauze et al. (1983), for the boundary conditions (i) $C_A = C_{A\infty}$, $\theta = 0$, (ii) $C_A = C_{Ao}$, $\theta > 0$, $r = d/2$, (iii) $C_A = C_{A\infty}$, $r = \infty$. It

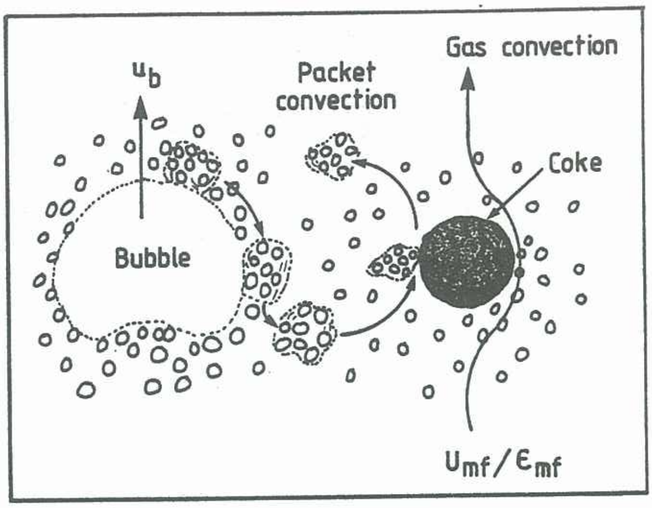

Fig. 2 Mechanism for oxygen transfer by particulate phase gas convection and packet transfer to a coke particle in a bubbling fluidized bed

may be shown that the mass transfer rate of A per unit surface area, F_A, is given by:

$$F_A = (C_{A\infty}-C_{Ao})\,\{\frac{2D}{d} + (\frac{4D}{\pi\theta_m})^{1/2}\} \qquad (6)$$

D is the diffusivity of A in the surrounding medium and θ_m is the contact time of the gas and packets at the surface.

Comparison of equation (6) with the conventional mass transfer relationship $F_A = \beta(C_{A\infty}-C_{Ao})$ provides:

$$\beta = 2\frac{D}{d} + (\frac{4D}{\pi\theta_m})^{1/2} \qquad (7)$$

or in terms of N_{Sh},

$$N_{Sh} = 2 + (\frac{4d^2}{\pi D\theta_m})^{1/2} \qquad (8)$$

At low Reynolds numbers in the absence of the fluidized bed it can be expected that for a given fluid θ_m is a function of (d/U). For a simple relationship θ_m = kd/U, where k is a constant or a slowly varying function of the flow conditions, it follows directly that:

$$N_{Sh} = 2 + (\frac{4}{\pi k})^{1/2}\,(N_{Re})^{1/2}\,(N_{Sc})^{1/2} \qquad (9)$$

Equation (9) is of the form of equation (1) but with a slightly different exponent for N_{Sc}. This lends theoretical support to the form of equation (1). Application of equation (7) to a fluidized bed requires an evaluation of the contact time, θ_m. La Nauze et al. (1983) proposed that the combined effect of the gas and packed contacts may be taken as the sum of the individual frequency of contact given by:

$$\frac{1}{\theta_m} = (\frac{U_{mf}}{d\varepsilon_{mf}} + \frac{u_b}{d}) \qquad (10)$$

where ε_{mf} is the voidage at U_{mf} and u_b is the absolute rising velocity of a bubble. It follows from equation (7), with the assumption that $D \simeq D_a\,\varepsilon_{mf}$ and $N_{Sh} = \beta d/D_a$, that:

$$N_{Sh} = 2\varepsilon_{mf} + \{\frac{4\varepsilon_{mf}\,d\,(U_{mf}/\varepsilon_{mf} + u_b)}{\pi D_a}\}^{1/2} \qquad (11)$$

The bubble velocity (u_b) in a fluidized bed may be evaluated from well established equations (Davidson et al., 1977).

5. EVALUATION OF N_{Sh} FOR BURNING PARTICLES

Sherwood numbers were determined for single spherical petroleum coke particles burning in air in fluidized beds of sand in a 102 mm diameter combustor. Details of the apparatus and techniques have been given by La Nauze and Jung (1982). The change in mass and diameter of a particle was determined for set periods of time. A typical initial mass was 1.2 g and diameter 12 mm; burn-out was followed for about 30 min, providing about 30 data points per particle. The combustion rate was determined from the mass loss per unit time.

The experimental data presented in the figures represents the change in N_{Sh} with burn-off of a single particle. The variability in the data is attributable in the early stages of burn-off to some "edge" rounding but is largely a result of an accumulation of errors in the mass and diameter measurements used to obtain the N_{Sh} values. A typical result derived from the combustion rate and particle temperature data is given in Figure 1; and for other operating conditions in the papers by Jung and La Nauze (1983) and La Nauze et al. (1983). Further experimental results are reported below.

6. COMPARISON OF EXPERIMENT WITH THEORY

6.1 Variation of N_{Sh} with d

Favourable comparison between theory (equation (11)) and experiment for the variation of N_{Sh} with d is shown in Figure 1. La Nauze et al. (1983) reported similar agreement over a range in coke particle diameter from 1.5 to 13 mm, inert particle diameter from 0.655 to 0.925 mm, bed temperature from 973 to 1173 K and fluidizing velocities up to 0.63 m/s. This suggests that for particle Reynolds numbers between 2 and 55, equation (11) satisfactorily predicts the mass transfer of oxygen to a large particle burning in a fluidized bed of smaller inert particles.

In Figure 1, the semi-empirical relationship for N_{Sh} (equation (2)) is also given. Equation (2) is almost identical to the predictions given by equation (11). This arises, as shown elsewhere (La Nauze et al., 1983) because $N_{Sc}^{1/3}$ and $(U_{mf}/\varepsilon_{mf} + u_b)/U$ are both approximately 1 for the conditions of the experiments. The derivation of the theory suggests that equation (2) should be modified so that 2ε is replaced by $2\varepsilon_{mf}$.

6.2 Variation of N_{Sh} with U

Figure 3 illustrates the variation in N_{Sh} with velocity. Since the diameter could not be held constant throughout an experiment, the range of N_{Sh} values obtained at a given value of d from several experiments is plotted. Equation (11) again provides a satisfactory fit to the data.

6.3 N_{Sh} for Packed Beds

Equation (11) reduces to an equation which could be applied to a packed (non-fluidized) bed when $u_b = 0$, viz:

$$N_{Sh} = 2\varepsilon + \left(\frac{4dU}{\pi D_a}\right)^{1/2} \qquad 0 < U < U_{mf} \qquad (12)$$

A number of experiments were performed in which combustion occurred with the sand bed packed around the burning particle. An example of the results is given in Figure 4, where reasonable comparison with equation (12) can also be seen.

7. CONCLUSIONS

A theory for mass transfer of a gas to a large particle in a fluidized bed of smaller non-reacting particles has been developed. It is proposed that mass transfer occurs by non-steady state diffusion from (i) the gas flow around the particle and (ii) packets of particles approaching the surface carrying fresh gas.

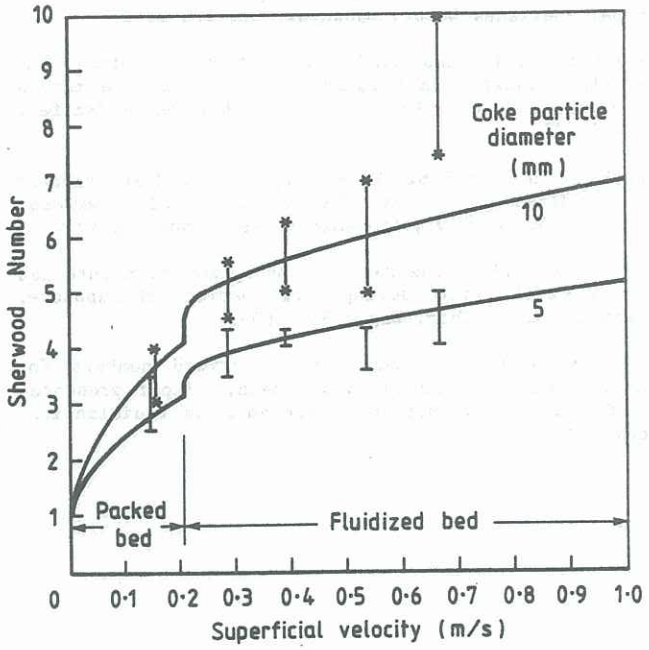

Fig. 3 Variation of N_{Sh} with superficial velocity for single petroleum coke particles of 5 and 10 mm diameter in air-fluidized beds of 0.78 mm diameter sand at 1173 K compared with variation predicted by equation (11). Ranges of experimental values are shown by bars; predicted variation by lines

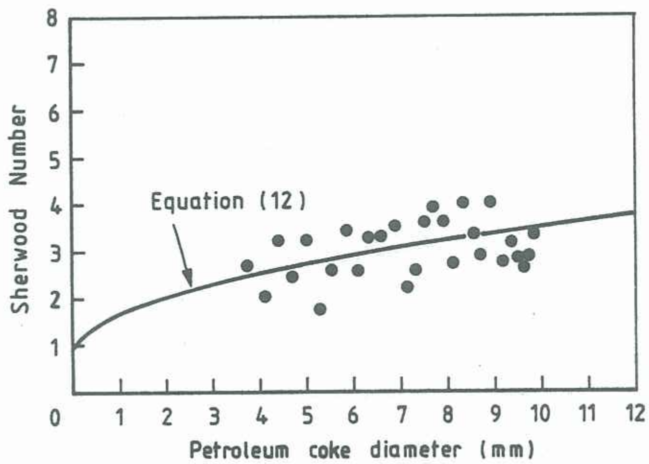

Fig. 4 Variation of N_{Sh} with d in a packed bed of 0.925 mm diameter sand at 1173 K and superficial air velocity 0.15 m/s. $\varepsilon = 0.39$

The theory is shown to predict the diffusion of oxygen to the particle for the combustion of single petroleum coke particles in fluidized and packed beds.

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