Evaporation in a Packed Bed Heated from the Bottom at Reduced Pressure

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Experiments were performed to measure the evaporation rate of water in a packed bed of equal-sized and wetted spherical particles when the packed bed was heated from the bottom at a reduced pressure. The evaporation rate based on an equivalent channel model with laminar countercurrent flow was first applied to the system, and the theoretical evaporation rate obtained was compared with the evaporation rate observed in the system.

THEORETICAL EVAPORATION RATE OF EQUIVALENT CHANNEL MODEL

Among many models of granular beds, the simple equivalent channel model was applied to the packed bed in this study.

The porosity of a packed bed was approximated as a perpendicular capillary tube, of which the radius is calculated for the equivalent channel model. It is assumed that the liquid is evenly distributed in the bed and flows down along the tube by gravitational force while the vapor rises as laminar flow without accompanying condensation. Subsequently, we applied the analytical results for a flow model in a wetted wall column by Hikita et al. (1976), who expressed the average velocities of flowing-down liquid and the uprising gas in laminar flow as follows:

$$U_{Lm} = \frac{\Delta P_f}{8\mu_L H}(R^2 - a^2) - \frac{(\rho_L - \rho_G)g}{8\mu_L}\left\{R^2 - 3a^2 - \frac{2a^4}{R^2 - a^2} ln\left(\frac{a}{R}\right)^2\right\} \quad (1)$$

$$U_{Gm} = \frac{\Delta P_f}{8H}\left\{\frac{2(R^2 - a^2)}{\mu_L} + \frac{a^2}{\mu_G}\right\} - \frac{(\rho_L - \rho_G)g}{4\mu_L} \times \left\{R^2 - a^2 + a^2 ln\left(\frac{a}{R}\right)^2\right\} \quad (2)$$

where

g = gravitational acceleration
H = height of packed bed
ΔP_f = frictional pressure drop for gas stream across a packed bed
μ_G, μ_L = viscosities of gas and liquid
ρ_G, ρ_L = densities of gas and liquid

The ratio of the voidage which is occupied by water to the whole voidage is defined as the moisture content ϕ.

The relationship among the moisture content ϕ, the equivalent channel radius R and the gas passing equivalent channel radius a is represented by $\phi = 1 - (a/R)^2$. Then, from Eqs. 1 and 2, U_{Gm} is expressed by Eq. 3.

$$U_{Gm} = \left\{2 + \frac{\mu_L(1-\phi)}{\mu_G\phi}\right\} U_{Lm} - \frac{gR^2(\rho_L - \rho_G)(1-\phi)}{8\phi} \times \left[\frac{2}{\mu_L}\left\{2\phi + (2-\phi)\, ln\,(1-\phi)\right\} - \frac{1}{\mu_G}\left\{3\phi - 2 - \frac{2(1-\phi)^2 ln\,(1-\phi)}{\phi}\right\}\right] \quad (3)$$

The gas mass flow rate W_{Gm} and the liquid mass flow rate W_{Lm} per unit cross-sectional area of the packed bed are related with U_{Gm} and U_{Lm} by use of A and B, and the following equation is obtained from Eq. 3.

$$W_{Gm} = \frac{\rho_G(1-\phi)}{\rho_L\phi}\left\{2 + \frac{\mu_L(1-\phi)}{\mu_G\phi}\right\} \times W_{Lm} - \frac{\pi A^4 r_p{}^2 g\rho_G(\rho_L - \rho_G)(1-\phi)^2}{8B\phi} \times \left[\frac{1}{\mu_G}\left\{3\phi - 2 - \frac{2(1-\phi)^2 ln\,(1-\phi)}{\phi}\right\} - \frac{2}{\mu_L}\{2\phi + (2-\phi)\, ln\,(1-\phi)\}\right] \quad (4)$$

Both A and B in Eq. 4 are constants, defined by the packing system: A refers to R/r_P and B refers to πR^2/(the area of an unit cell).

The flooding takes place with an increasing U_{Gm}. Since liquid flows down so slowly under this state, the evaporation rate at the bottom decreases. Therefore, the maximum flow rate of the upward gas is equal to the evaporation rate at the flooding point. The gas flow rate W_{Gf} at flooding point is obtained from Eq. 4 by substituting $W_{Lm} = 0$. If the bottom of the packed bed is heated effectively, flooding takes place intermittently. Therefore, the evaporation rate can be approximated by W_{Gf}, which is the product of Y_f and Z.

$$Z = \pi A^4 r_P{}^2/B \quad (5)$$

$$Y_f = \frac{1}{8}g\rho_G(\rho_L - \rho_G)\frac{(1-\phi)^2}{\phi}\left[\frac{1}{\mu_G}\left\{3\phi - 2 - \frac{2(1-\phi)^2 ln\,(1-\phi)}{\phi}\right\} - \frac{2}{\mu_L}\{2\phi + (2-\phi)\, ln\,(1-\phi)\}\right] \quad (6)$$

The values of A and B were obtained by our previous geometric calculation (1978).

The average value of $Y_f, \bar{Y}_f$ is obtained by integrating Eq. 6 from ϕ_1 to ϕ_2.

On the other hand, the critical condition for laminar flow has to be confirmed. The gas Reynolds number at the flooding point might be expressed by using the equivalent channel radius and the gas velocity in the equivalent channel as the representative length and the velocity; i.e.,

$$Re_{Gf} = 2A^3 r_P{}^3 Y_f/(\mu_G\sqrt{1-\phi}) \quad (7)$$

According to the Blake-Kozeny's equation that expresses the flow rate in a packed bed of equal-sized spherical particles with the hydraulic radius, the condition for laminar flow in terms of Reynolds number is given by Bird et al. (1960) as:

$$2r_P U_o \rho_G/\{\mu_G(1-\epsilon)\} < 10 \quad (8)$$

Where, U_o and ϵ mean the average superficial gas velocity and the porosity of packed bed, respectively. By using U_o for velocity, presentative diameter of flow path and tortuosity which are used in the Blake-Kozeny's equation, the following equation is obtained from Eq. 8.

$$Re_{Gf} < 10A(1-\epsilon)\sqrt{1-\phi} \quad (9)$$

EXPERIMENTAL APPARATUS AND PROCEDURE

The evaporation vessel of the experimental apparatus for this study consists of a double tube made of acrylic resin. The diameter of the inner tube was 12 cm and that of the outer tube 15 cm. The annular space between the two tubes was held at a reduced pressure to avoid heat transfer.

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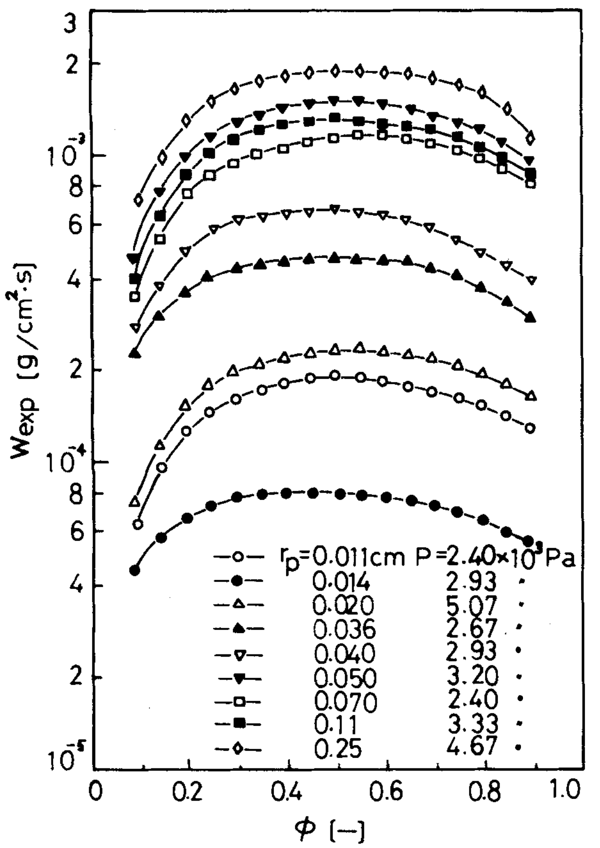

Figure 1. Evaporation rate.

The inner tube was packed with glass beads up to 20 cm from the bottom and filled with water enough to just immerse the beads. The particle diameters of glass beads were measured by a magnifying viewer after sieving them. The bottom of the vessel was made of a copper plate which functioned as a part of the steam jacket. Eight thermocouple sensors were arranged perpendicularly in equal distance at 2 cm apart from the center of the tube, and six dielectric hygrometer sensors were similarly arranged but at 4 cm apart from the center. The evaporation rate was measured by collecting condensed vapor in a burette. In this measurement, the loss due to the vacuum pump was calibrated. The Hg-manometer was mainly used to measure the operating pressure P at the top of the bed. Pressures, temperature, moisture content, and the volume of condensed water were measured at interval of five minutes.

EXPERIMENTAL RESULTS AND CONCLUSIONS

A typical evaporation rate W_{exp} observed in the moisture content between 0.1 and 0.9 is shown in Figure 1.

The evaporation rate increased in proportion to r_P and P, and the maximum value was obtained at $\phi = 0.5$.

In this study, the random packing system was used, and the porosity ϵ of packed bed was found to be 0.385. This was almost the same as the theoretical porosity 0.3954 of the regular tetragonal or hexagonal orthorhombic packing system. Then, the equivalent channel radius of the random packing system was assumed to be equal to that of the regular packing system. By our previous calculation (1978), the mean radius $\bar{R}$ obtained was $0.612 r_P$, hence A = 0.612, B = 2.866. Y_{exp} was obtained by dividing W_{exp} by Z in Eq. 5.

In order to appraise the proximity of Y_{exp} and Y_f under different operating pressures, the arithmetic means $\bar{Y}_{exp}$ over the moisture content range 0.1 through 0.9 was used. Then, $\bar{Y}_f$ was calculated by Eq. 6 and compared with $\bar{Y}_{exp}$ in Figure 2.

As can be seen from Figure 2, the discrepancy of $\bar{Y}_{exp}$ from $\bar{Y}_f$ is large only when r_P is very small or large. When it is very small, the adhesion force due to surface tension exceeds the gravitational force and the evaporation is apt to accompany with bumping or channeling.

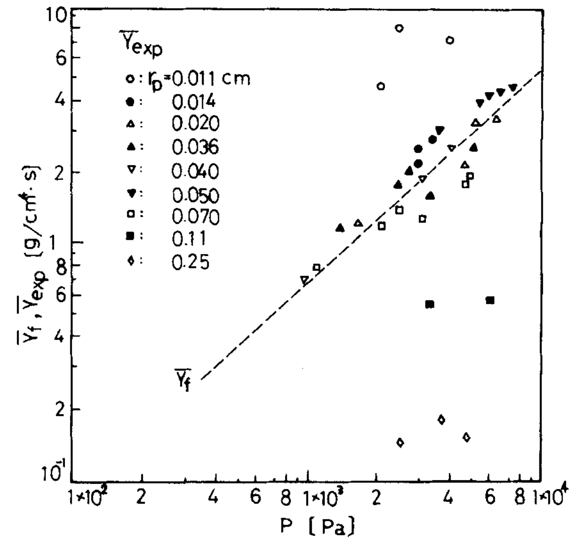

Figure 2. Correlation between the mean theoretical value ($\bar{Y}_f$) and the mean observed value ($\bar{Y}_{exp}$) of the gas flow velocity.

When r_P is very large, the vapor flow is supposed to become turbulent. By applying to Eq. 9, $A = 0.612$, $\epsilon = 0.385$, and $\phi = 0.36$ at the maximum theoretical evaporation rate, $Re_{Gf} < 3.0$ as the condition necessary for laminar flow is obtained.

When r_P is larger than 0.11 cm, the value of Re_{Gf} calculated from Eq. 7 is larger than 3.0. This supports the assumption that the flow pattern of gas is not laminar but turbulent when r_P is larger than 0.11 cm. Consequently, the basic assumption for the evaporation model is not held in the above cases.

At operating pressures between 9×10^2 and 8×10^3 Pa, and $0.014 < r_P < 0.07$ cm, a good agreement was found between $\bar{Y}_f$ and $\bar{Y}_{exp}$. Therefore, the present model may be applied to this experiment system under these conditions.

NOTATION

A	= constant, defined by the packing system, R/r_p
a	= gas passing equivalent channel radius (cm)
B	= constant, defined by the packing system, πR^2/(the area of a unit cell)
g	= gravitational acceleration (cm/s^2)
H	= height of packed bed (cm)
P	= operating pressure (Pa)
R	= equivalent channel radius (cm)
$\bar{R}$	= mean equivalent channel radius (cm)
Re_{Gf}	= gas Reynolds number at flooding point
r_p	= particle radius (cm)
U_{Gm}, U_{Lm}	= flow velocities of gas and liquid (cm/s)
U_o	= average superficial gas velocity (cm/s)
W_{exp}	= observed evaporation rate (g/cm^2·s)
W_{Gf}	= gas flow rate at flooding point (g/cm^2·s)
W_{Gm}, W_{Lm}	= flow rate of gas and liquid (g/cm^2·s)
Y_{exp}	= observed relative evaporation rate (g/cm^4·s)
$\bar{Y}_{exp}$	= mean observed relative evaporation rate (g/cm^4·s)
Y_f	= relative gas flow rate at flooding point (g/cm^4·s)
$\bar{Y}_f$	= mean value of Y_f, between ϕ_1 through ϕ_2 (g/cm^4·s)
Z	= characteristic value of packed bed (cm^2)
ϵ	= porosity of packed bed
ΔP_f	= frictional pressure drop (Pa/cm)
μ_G, μ_L	= viscosities of gas and liquid (Pa·s)
ρ_G, ρ_L	= densities of gas and liquid (g/cm^3)
ϕ	= relative moisture content

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Coal Particle Suspensions in Vertical Downflow

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Entrained flow reactors, where particles suspended in a gas flow cocurrently through a heated tube, are of considerable interest in coal conversion processes due to their high throughputs. The verticle downward flow orientation (downflow) is of particular interest in such reactors (as compared with horizontal flow or upflow) because of lower total pressure drop and no minimum gas flow rate. Unfortunately, general design correlations for predicting pressure drop and heat transfer characteristics of flowing gas-solids suspensions do not exist for any tube orientation, even for suspensions of uniform, spherical particles. Moreover, data for the downflow orientation are surprisingly scarce, and upflow data are not generally applicable to downflow, as has been shown by Kim and Seader (1983a,b).

Those authors investigated pressure drop and heat transfer characteristics of a suspension of uniform 329-μm glass beads in air, flowing cocurrently downward through a vertical 0.0180-m-i.d. tube. They found the frictional pressure drop to increase with solids loading at a rate significantly lower than is typically reported in the literature for upflow, and the wall Nusselt number to be nearly independent of solids loading. They concluded that transport mechanisms for flowing gas-solid suspensions might be generally weaker in downflow than in upflow.

In this study, data were obtained at similar conditions using suspensions of coal particles. Pressure drop, particle velocity, particle Nusselt number, and wall Nusselt number were determined for fully developed flow, for average particle sizes of 100 and 300 μm. Gas Reynolds number varied from 10,000 to 30,000, and solids loading varied from 0 to 20 kg coal/kg air.

The experimental apparatus, as described by Brewster (1979), consisted of a 7.3-m length of 0.0126-m-i.d. s.s. tubing which was instrumented with 14 pressure taps, and two photomultiplier tubes for determining the velocity of injected particles tagged with phosphorescent powder as described by Brewster and Seader (1980). For heat transfer measurements, a 1.29-m length of tubing at the lower end of the test section was replaced with an identical section which was instrumented with 21 thermocouples for measuring the axial wall temperature profile. The heat transfer section was heated directly by a DC current. In order to insure a fairly constant heat loss for all runs, the maximum wall temperature (near the lower end of the heat transfer section) was adjusted to 366.5 ± 1.1 K, by manually varying the rheostat on the welding unit. This temperature is low enough that significant changes in the coal due to heating were avoided.

A thermocouple sheathed in a s.s. tube was inserted into the suspension at the lower end of the heat transfer section for measuring exit gas temperature. The method of accounting for particle impact-heating and determining the gas temperature has been described by Brewster and Seader (1983). Since this thermocouple quickly eroded, outlet gas temperature was measured for only a few runs.

After checking the system and determining the heat loss with air alone, complete pressure drop profiles were obtained with the larger coal particles to determine whether the test section was long enough to achieve fully developed flow. The profiles appeared asymptotic at the lower end (Brewster, 1979), thus suggesting that the flow was indeed fully developed.

Frictional pressure drop for the region of fully developed flow was calculated from the total measured pressure drop by subtracting the negative contribution due to the solids static head. The *in-situ* concentration of particles was calculated assuming no slip between the particles and gas. This assumption was supported by the results of the particle velocity measurements (particle velocity agreed with gas velocity within experimental error) and by a sensitivity analysis, which showed that the assumption of a slip velocity equal to the free-particle settling velocity does not significantly affect the results.

Frictional pressure drop was correlated in terms of specific pressure drop, defined as the ratio of the frictional pressure drop with solids to that of gas alone flowing at the same velocity. As shown in Figure 1, the frictional pressure drop always increases with increasing solids loading, but the rate of increase is very slow, even slower than that reported by Kim and Seader for glass beads of roughly comparable size. This difference is probably due to entrance effects, since glass beads require a longer acceleration length than coal particles due to higher density and spherical shape.

Particle Nusselt number was determined by heat balance, using the following one-dimensional thermal energy equations for gas and solids:

$$T_p(x) = T_p(0) + \frac{\alpha\beta}{\alpha+\gamma}x - \frac{\alpha\beta}{(\alpha+\gamma)^2}[1 - e^{-(\alpha+\gamma)x}] \quad (1)$$

$$T_g(x) = T_g(0) + \frac{\alpha\beta}{\alpha+\gamma}x + \frac{\beta\gamma}{(\alpha+\gamma)^2}[1 - e^{-(\alpha+\gamma)x}], \quad (2)$$

where α, γ, and β depend on particle Nusselt number Nu_p, particle

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