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II. Experimental Evidence for Axial Dispersion and Intraparticle Effects

The effects of axial fluids dispersion on breakthrough curves in packed beds is considered and formulae are derived for 'F-factors', which allow heat transfer coefficients obtained by step testing to be corrected for axial fluids dispersion. A comparison of steady state and unsteady correlations for heat transfer in packed beds suggests that axial dispersion effects are incorporated in certain steady state heat transfer correlations. Further correction would thus lead to errors in breakthrough curve prediction. Experimental data for breakthrough curves in packed beds of highly conductive and poorly conductive spheres is examined and Biot number correction factor $1 + Bi/5$ shown to be valid for Biot numbers up to about 4.

Part I of this paper (*8*) presents simplified theoretical methods of predicting breakthrough curves in linear adsorbers and thermal regenerators, allowing for axial fluids dispersion. Unfortunately, experimental evidence for the effects of axial fluids dispersion on packed bed dynamics is inconclusive and difficult to interpret. Reasons for this are examined in this paper.

Particle internal conduction effects are more easily predicted and measured. Experimental evidence for the validity of the $1 + Bi/5$ correction factor suggested by Bradshaw et al. (*12*) and discussed in Part I of the paper is examined using the extensive results of Handley and Heggs (*7*) together with some confirmatory results of the writer.

AXIAL DISPERSION EFFECTS—EXPERIMENTAL DIFFICULTIES

The integral deviation square computations of Part I (*8*) suggest that experimentally the effects of axial fluids dispersion may be difficult to separate from that due to fluid-particle heat transfer resistance in the analysis of breakthrough curves in packed beds. This is because the predicted deviations between various models of packed bed dynamics due to axial fluids dispersion are generally less than would arise from experimental errors. As an example, Table 1 shows the integral deviation square between impulse responses predicted by the Schumann model, (which ignores axial fluid dispersion) and the continuous, axial dispersion model with simplified boundary conditions of Chao and Hoelscher (*13*) discussed in Part I of this paper. The comparison is at a fixed normalized variance σ^{*2} of 0.125, a heat capacity ratio V_H of 10^4 and a Biot number of zero.

An integral deviation square of 5.4×10^{-4} (obtained when axial dispersion contributes 40% to total variance σ^{*2}) would be extremely difficult to distinguish from experimental 'noise' and other inaccuracies.

Nonetheless, the effects of axial dispersion could be distinguished from fluid-solid heat transfer if an accurate heat transfer coefficient correlation (obtained by methods which are free from axial dispersion) were available. Any increase in breakthrough curve dispersion over that pre-

Table 1. Integral Deviation Squares Between Schumann and Axial Deodersion Models for Normalized Variance of 0.125, $Bi = 0$ and $V_H = 10^4$

NPe	Axial dispersion contribution to variance, %*	Y	Integral deviation square
2000	0.8	16	1.5×10^{-8}
160	10	17.8	2.3×10^{-6}
80	20	20.0	3.4×10^{-5}
40	40	26.6	5.4×10^{-4}

* According to Equation (15) Part I of this paper (8).

dicted by the fluid-particle heat transfer contribution would be due, in the absence of intraparticle diffusion and axial 'fluid-solid-fluid' conduction, to axial fluids dispersion. This is the approach used by Babcock and coworkers (*14*) who employed the mass transfer correlation of Dryden (*15*) assuming j_H factors for heat and mass transfer to be equal.

INDEPENDENT HEAT TRANSFER CORRELATIONS

Unsteady state measurements of heat transfer coefficient cannot yield the independent correlation required. A detailed literature survey was carried out on the various steady state measurements of heat transfer coefficients in packed beds satisfying the constraints:

tube to particle diameter ratio $\geqq 15$
packing: random, $0.35 \leqq \epsilon \leqq 0.40$, where ϵ = porosity,
bed length to particle diameter ratio, $N > 3$
bed geometry: circular
particle shape: spherical

Coefficients averaged over entire bed were required, instead of point coefficients.

These criteria eliminated all except three of the correlations reported in Barkers extensive survey of heat transfer in packed beds (*1*) of which one, Baumeister and Bennett's (*2*), was rejected because of the method used (induction heating of steel spheres) resulted in large radial temperature differences in the bed. The two remaining, those of Denton (*4*) and Bradshaw and Meyers (*3*) involved, respectively, measurements of heat input and temperature differences for isolated spheres in a packed bed and steady state evaporation from porous spheres. At longer bed lengths Bradshaw and Meyers found it necessary to numerically integrate to obtain the 'true' mean temperature differences. This would be necessary if axial fluid dispersion were important although this is unlikely at Reynolds number greater than 400 as in their work. Because of large experimental errors inherent in the porous spheres approach (explained by Bradshaw and Meyers) greater weight is given to Denton's work for the present purposes.

F FACTOR APPROACH TO AXIAL DISPERSION

The predicted effect of axial fluids dispersion on the 'spread' σ^* of a breakthrough curve (*8*) may be obtained by use of Equation (15) of Part I of this paper (assuming a Biot number of zero and negligible axial fluid-solid-fluid conduction):

$$\sigma^{*2} = \frac{2}{NPe} + \frac{2}{Y} \cdot \beta^2 = \frac{2}{Y_{\text{calc}}} \cdot \beta^2 \tag{1}$$

In Equation (1) as before Y_{calc} is the apparent heat transfer factor obtained if axial dispersion is assumed absent and β is defined in this part of the paper as the ratio $V_H/(V_H + 1)$.

The equivalent Peclet number can only be obtained with accuracy from an experiment which yields σ^{*2} (the normalized variance of the impulse or chromatographic response) when the ratio P = fractional contribution of axial dispersion to total variance is reasonably high. Defining a further ratio

$$F = \frac{St_{\text{calc}}}{St} = \frac{Y_{\text{calc}}}{Y} \tag{2}$$

we obtain by manipulation of Equation (1)

$$F = 1 - P = \frac{\beta^2 NPe}{\beta^2 NPe + Y} = 1 - \frac{Y_{\text{calc}}}{\beta^2 NPe} \tag{3}$$

Equations (3) show that the relative effect of axial dispersion depends only on the Peclet number Pe, the Stanton number St, and bed porosity and is independent of bed length for a given heat capacity ratio, that is,

$$F = \frac{\beta^2 Pe}{\beta^2\, Pe + 6\, St(1 - \epsilon)} = 1 - \frac{6\, St_{\text{calc}}}{\beta^2 Pe} \cdot (1 - \epsilon) \tag{4}$$

F factors become smaller as the ratio of the Stanton to the effective Peclet number becomes larger and this occurs at lower flowrates. At very low flowrates, axial fluid-solid-fluid conduction may be expected to cause additional spreading of a breakthrough curve.

COMPARISON WITH STEADY STATE F FACTORS

Epstein (*6*) has derived equations for F factors which allow true overall or integrated mean temperature differences ΔT to be estimated from the log mean temperature difference ΔT_{LM} in experiments involving heat transfer to spheres of constant surface temperature (assuming the validity of a finite stage or mixing model in such experiments) where in Epstein's work

$$\Delta T = F \cdot \Delta T_{LM} \tag{5}$$

Epstein's definition of F is identical to that of Equation (2) if, for Epstein's porous sphere experiments, St_{calc} is the apparent Stanton number obtained by ignoring axial dispersion

$$St_{\text{calc}} \cdot \Delta T_{LM} = St \cdot (F\, \Delta T_{LM}) = (St \cdot F) \cdot \Delta T_{LM}$$

Epstein concluded that F approaches unity as the bed length N becomes large. This is not necessarily so, and it is shown in (*9*) that F factors are also independent of bed length in Epstein's case.

It should be noted that definition of F factors by Equation (2) requires that solutions obtained with and without axial dispersion have an *identical shape*. In the case of steady state heat transfer to spheres of uniform surface temperature this requires an exponential relation between fluid temperature and bed length with and without axial dispersion. Definition of the corresponding unsteady state F factors require that predicted breakthrough curves with and without axial dispersion be identical. As shown in Part I of the paper, this is only approximately so. Comparison of the F factors obtained from Equation (3) with the steady state factors derived in (*9*) for heat transfer to spheres of uniform surface temperature according to the axial disper-

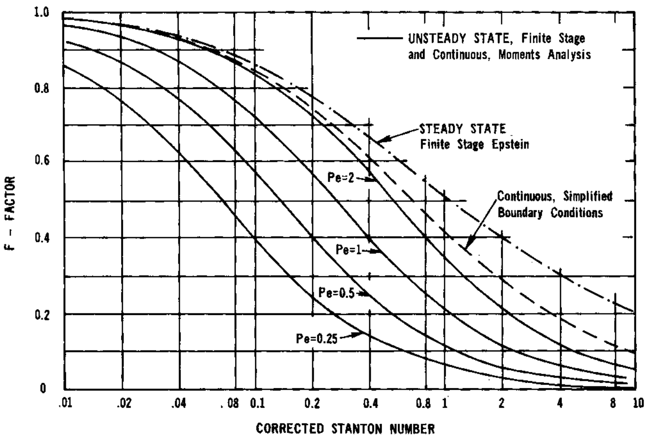

Fig. 1. *F* factors for steady and unsteady state heat transfer in packed beds.

sion model with simplified boundary conditions

$$F = \frac{Pe}{12\ St\ (1-\epsilon)} \left[1 + \frac{24\ St\ (1-\epsilon)}{Pe} \right]^{1/2} - 1 \qquad (6)$$

yields very similar values of F (Figure 1). As before, F is independent of bed length although the use of the full Danckwerts boundary conditions causes a small length-dependent term to be added to Equation (6).

CORRECTION OF HEAT TRANSFER CORRELATIONS FOR AXIAL DISPERSION

Handley and Heggs (7) have obtained a heat transfer correlation by fitting the Schumann step response solution to experimental breakthrough curves. It will be recalled that the Schumann model makes no allowance for axial fluid dispersion. The Handley and Heggs correlation of modified j_H factor against superficial Reynolds number is for spherical metallic packings (steel, lead, bronze):

$$j_H \cdot Re_p \cdot \epsilon = 0.255\ Re_p^{0.665} \qquad (7)$$

or for a voidage ϵ of 0.37 as in Denton's experiments and a Prandtl number of 0.74

$$St = 0.845/Re_p^{0.335} \qquad (8)$$

Regarding the Stanton numbers calculated from Equation (7) as apparent coefficients St_{calc} Equation (3) may be used to obtain F factors and the fraction P of total variance resulting from axial dispersion. The results of a few calculations are given below in Table 2.

It will be noted that the axial dispersion correction is about 34% at the low flow ($Re_p = 100$) falling to 10% at the higher Reynolds number of 4,000. The correction does not fully account for the deviation between the Handley and Heggs' work and Denton's. No special attempts were made to reduce spreading of the input step in Handley and Heggs' work, apart from the use of a fine wire electrical air heater. The presence of significant spreading of the input step, usually noticeable at higher flowrates would cause an increased value of output curve spread σ^* to occur which would in turn result in an apparently lower value of Stanton number.

A number of step tests have been carried out on a bed of spheres taking precautions to eliminate the effects of the thermal inertia of the 0.010 in. 'Nichrome' heater wires used to administer the inlet temperature changes (*10*).

Experimental conditions are listed in Table 3. Data analysis was by the method suggested in Part I of this paper, that is, time differences between 5 and 80%, 10 and 80%, 20 and 80%, and 40 and 80% of correlated response were first normalized by dividing by $\mu(t)$, the first moment of the impulse response and the resultant values of ΔT^* then used via the empirical correlations given in Part I to estimate the normalized variance σ^{*2} of the impulse response. Values of St_{calc}, the apparent Stanton number, uncorrected for axial fluids dispersion were then calculated from Equation (1) and correlated against superficial Reynolds number. Estimates of σ^{*2} from the four time differences did not differ by more than 2% from the mean for a particular step test.

As may be seen from Figure 2, the uncorrected Stanton numbers give excellent agreement when compared with the Denton correlation. The result of correcting, for axial

TABLE 2. AXIAL DISPERSION CORRECTIONS TO DATA OF HANDLEY AND HEGGS
$\epsilon = 0.37, \quad Pe = 2$

				Stanton numbers	
Re_p	St_{calc} Equation (7)	P, %	F Equation (4)	Corrected, Handley and Heggs	Denton
100	0.1808	34.2	0.658	0.275	—
200	0.1432	27.1	0.729	0.197	—
400	0.1136	21.4	0.786	0.144	0.1197
800	0.0901	17.0	0.830	0.108	0.0972
1000	0.0835	15.8	0.842	0.099	0.0909
2000	0.0665	12.6	0.874	0.076	0.0738
4000	0.0528	10.0	0.900	0.059	0.0600

TABLE 3. EXPERIMENTAL CONDITIONS FOR LEAD PACKING STEP TESTS

Packing: 6.3 in. of 4.04 mm. 'ICI' lead shot and 0.3 in. of 6.2 mm. 'Englass' soda glass
Weight of lead: 17.09 lb.
Weight of glass: 0.24 lb.
Specific gravity of lead: 11.2 at 65°F.
Approximate heat capacity of lead: .032 B.t.u./(lb. °F.)

Porosity: 0.361 (random packing)
Bed diameter inside PVC insulation: 3.77 in.
Tube to particle diameter ratio: 23.6
Pressure: 5.0 lb./sq.in.gauge
Thermocouples in thermopile across the section surrounded by soda glass insulation
Inlet air temperatures ranged from 65° to 80°F. with maximum inlet change of 20°F.
More details of these tests are available in (9).

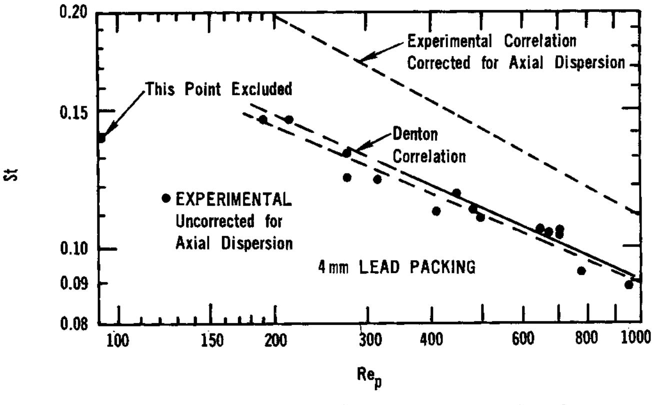

Fig. 2. Experimental Stanton number correlation, lead packing.

dispersion, a least squares fit through the St_{calc} versus Re_p data is also shown on Figure 2—the corrections range from 27% at $Re_p = 200$ to 17% at a Reynolds number of 1,000.

DISCUSSION OF AXIAL DISPERSION RESULTS

Equation (4) allows the theoretical contribution of axial fluids dispersion to the spread of a breakthrough curve to be evaluated rapidly given either the apparent Stanton number St_{calc} or the actual number St and an estimate of the fluid dispersion Peclet number. Theoretically, the contribution of axial dispersion will become very large at low Reynolds numbers if (as predicted for an isolated sphere in an infinite fluid) the Nusselt number approaches a constant value of approximately 2. In this event, since $St = Nu/(Re_p \cdot Pr)$, the Stanton number becomes very large, while the fall off in Peclet number predicted by Edwards and Richardson (5) will further exaggerate fluid dispersion effects. For heat transfer at any rate, Littman, Barrille, and Pulsifer (*11*) have shown that the picture is further complicated by the effects of fluid-solid-fluid axial conductivity for Reynolds numbers less than 100.

The absence of a reliable heat transfer correlation which can be conclusively shown to be free from axial dispersion effects at Reynolds numbers less than 500 makes separation of the two effects extremely difficult since as pointed out before, axial dispersion has a very small effect on the shape of an impulse or step response. Even Denton's correlation cannot be free of axial dispersion effects since axial dispersion must affect the 'true' average temperature difference around an isolated sphere in a bed.

Agreement between the apparent (uncorrected) Stanton numbers St_{calc} obtained from the present lead packing step tests and Denton's correlation suggests that from a practical viewpoint, axial dispersion and fluid-solid heat transfer may best be lumped into an apparent Stanton number St_{calc} especially for $Re_p > 500$. Hence, in the following discussion of 'intraparticle' or internal particle conduction effects, Equation (15) of Part I of the paper becomes

$$\sigma^{*2} = \frac{2}{Y_{\text{calc}}} (1 + Bi_{\text{calc}}/5) \cdot \beta^2 \qquad (9)$$

where Bi_{calc} is the apparent Biot number, uncorrected for axial dispersion. It will be noted that the bed length parameter, γx defined in Part I of the paper is still independent of axial dispersion, that is,

$$\gamma x = \frac{Y_{\text{calc}}}{B_{\text{calc}}} = \frac{Y}{Bi}$$

so that Equation (9) may also be written

$$\sigma^{*2} = \frac{2\beta^2}{Y_{\text{calc}}} + \frac{0.4\beta^2}{\gamma x} = \frac{2}{Y_p} \cdot \beta^2 \qquad (10)$$

where Y_p is a pseudo apparent heat transfer factor now incorporating both particle conductivity and axial dispersion effects.

PARTICLE CONDUCTION EFFECTS—EXPERIMENTS OF HANDLEY AND HEGGS

The experiments of Handley and Heggs on both highly and poorly conductive packings provide a useful source of data to test the supposed proportionality of the curve spread parameter σ^* and $\sqrt{(1 + Bi/5)}$ discussed in Part I of the paper. While it has been suggested above that spreading of the input step may have resulted in low values of their apparent Stanton number especially at higher flows, the error would presumably be present in experiments on both poorly and highly conductive packings to a similar extent for similar bed lengths and would be expected to cancel out.

Figure 3 illustrates the values of pseudo modified Colburn j_H factor obtained by Handley and Heggs for both metal spheres (where they are unaffected by particle conductivity) and soda glass of approximate thermal conductivity 0.60 B.t.u./(hr.) (ft.) (°F.). The pseudo glass sphere j_H factors were obtained by varying Y_p until a satisfactory fit between the soda glass step response and the Schumann solution resulted.

Then Equation (9) and Handley and Heggs' highly conductive particle correlation [Equation (7)] may be used to calculate theoretical values of Y_p for comparison with the soda glass experimental results.

Equation (9) then becomes

$$\frac{1}{St_p} = \frac{1}{St_{\text{calc}}} + Pr \cdot Re_p \cdot \frac{k_f}{10k_s}$$

Assuming $Pr = 0.74$ and $k_f = 0.0145$ B.t.u./(hr.) (ft.) (°F.) for air and $k_s = 0.60$ B.t.u./(hr.) (ft.) (°F.) for soda glass as suggested by Handley and Heggs (a value of 0.57 ± 8% was determined in work reported in (9)), we have

$$St_p = \frac{St_{\text{calc}}}{1 + St_{\text{calc}} \cdot Re_p/811} \qquad (11)$$

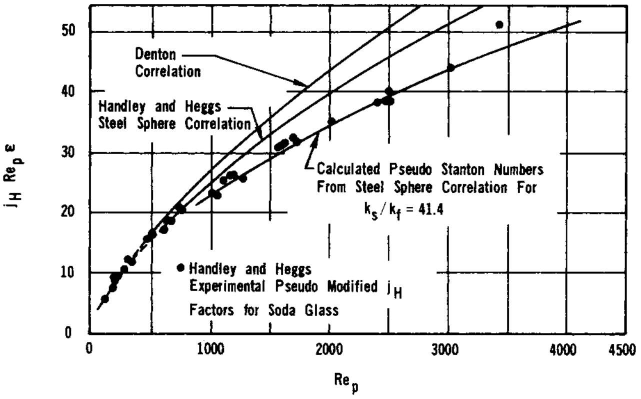

Fig. 3. Pseudo modified j_H factor of Handley and Heggs for soda glass.

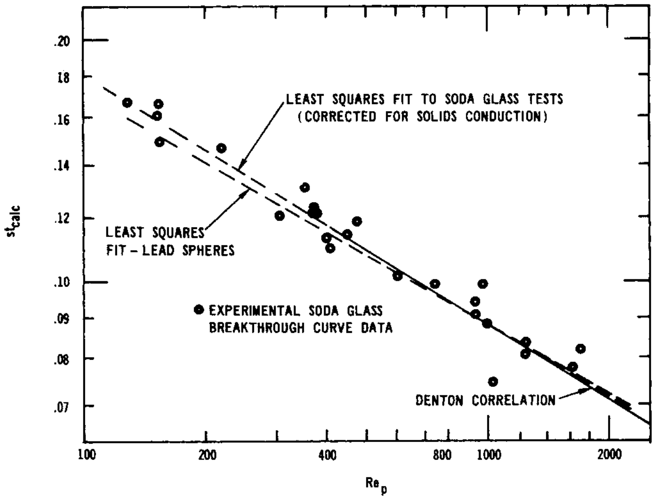

Fig. 4. Experimental Stanton number correlation, soda glass packing.

Table 4. Experimental Conditions for Soda Glass Packing Step Tests

Packing: 6.2 mm. 'Englass' soda glass spheres
'Bed lengths' or distance between inlet free stream thermocouples and immersed thermocouples in particle diameters: 3.30 in., and 12.50 in. for experiments reported in (9) and 3.30 in. for later experiments (when weight of packing was 2.18 lb.)
Specific gravity of packing: 2.583 at 77°F.
Volumetric thermal capacity of packing: 29.6 B.t.u./(cu.ft.)(°F.)

Porosity: 0.356 experiments reported in (9)
0.364, later experiments.
Thermal conductivity of melted slab of packing: 0.57 B.t.u./(Hr.)(ft.)(°F.) ± 8%
Bed diameter inside PVC liner: 3.77 in.
Tube to particle diameter ratio: 15.5
Pressures ranged from 5 to 50 lb./sq.in.gauge

where porosity is taken at 0.37. The calculated pseudo Stanton numbers from Equation (11) are superimposed onto Figure 3 and give excellent agreement with Handley and Heggs' experimental data, illustrating the good internal consistency of their results. At a superficial particle Reynolds number of 4,000, the Biot number uncorrected for axial dispersion would be about 3.6.

FURTHER EXPERIMENTS USING SODA GLASS PACKING

As a check on Handley and Heggs' results a further series of step tests on 6.2 mm. soda glass beds has been carried out and analyzed by the empirical correlation method of data analysis used for the 4 mm. lead packings. Experimental conditions are listed in Table 4. Values of St_{calc} thus obtained are plotted on Figure 4 and compared with the lead sphere correlation obtained before. Because of air flow limitations the maximum Biot number was about 1.5. Figure 4 includes an additional 10 runs not reported in (9).

The correlation obtained for $130 < Re_p < 1700$ was

$$St_{\text{calc}} = 0.76/Re_p^{0.31}$$

with 95% confidence limits of ± 12% and a correlation coefficient of −0.97.

When a t test was carried out to compare the regression coefficient b in $St_{\text{calc}} = a/Re_p^{b}$ with that obtained for the lead step tests, the value of t did not reach the 0.2 significance level. For 4 mm. lead

$$St_{\text{calc}} = 0.752/Re_p^{0.289}$$

with 95% confidence limits of ± 8.2% and correlation coefficient −0.96 where $230 \leq Re_p \leq 950$.

CONCLUSIONS

While the F factors derived in this paper do allow the theoretical effects of axial dispersion to be evaluated readily, it appears from close agreement between the uncorrected Stanton number correlations and Denton's data that the application of the corrections to apparently independent correlations will result in errors in prediction of breakthrough curves. It is suggested that fluid particle transfer correlations in fact may include their own axial dispersion corrections. It seems that the effects of axial 'dispersion' cannot be separated from those of fluid-particle transfer experimentally. The data of Handley and Heggs and the present experimental results support the validity of the $1 + Bi/5$ correction factor. Experiments at very high values of Biot number using (say) Perspex should be carried out to determine whether agreement continues for higher Biot number.

NOTATION

Bi, Bi_{calc} = fluid particle Biot number for heat transfer corrected and uncorrected for axial fluid dispersion, respectively, $Bi = Rh_p/k_s$
F = correction factor for axial dispersion, $St = St_{\text{calc}}/F$
h_p = surface-averaged fluid particle heat transfer coefficient, B.t.u./(hr.)(sq.ft.)(°F.)
k_s = thermal conductivity of solid packing, B.t.u./(hr.)(ft.)(°F.)
N = bed length in particle diameters
Pe = equivalent fluid phase axial dispersion Peclet number with particle diameter as characteristic dimension
R = particle radius, ft.
St, St_{calc} = fluid-particle Stanton number, corrected and uncorrected for axial dispersion, respectively
ΔT_{LM} = logarithmic mean temperature difference for fluid particle heat transfer, F°.
ΔT^* = normalized time difference on a breakthrough curve
V_H = ratio volumetric heat capacity B.t.u./°F. of packing to that of fluid
Y_p = pseudo heat transfer parameter, uncorrected for particle conductivity or axial dispersion
Y = heat transfer parameter $= 6NSt(1 - \epsilon)$ for spherical or right-cylindrical particles

Greek Letters

β = ratio, $V_H/(V_H + 1)$
ϵ = bulk volumetric porosity of bed
γx = 'bed length parameter' $= Y/Bi = Y_{\text{calc}}/Bi_{\text{calc}}$
σ^* = 'curve spread parameter' or standard deviation of impulse response of packed bed normalized relative to the stoichiometric breakthrough time

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