ELUTRIATION OF FINES FROM FLUIDIZED BED*

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The elutriation rate of fine particles was measured in the various systems of fluidized beds; gas-solid and liquid-solid, and continuous and batch fluidized systems.

From experiments on the continuous fluidized bed with two particle-size system, that is, fine and coarse particles, it appeared that the concentration of fines was kept at a constant value C_s throughout the bed height except in the space near the overflow pipe.

The elutriation rate coefficients of fines were calculated based on C_s .

Correlating the elutriation rate coefficients with the dimensionless groups introduced from the material and momentum balance equations, an empirical equation has been derived within $\pm 25\%$ deviation.

Introduction

It is important to study elutriation of fine particles from fluidized beds because it affects residence time, conversion and so on in drying, classifying and reaction by fluidized beds.

Leva¹⁾, and Osberg and Charlesworth²⁾ measured elutriation of fines from a batch fluidized bed of fine and coarse particle system (two particle-size system). Leva¹⁾ has proposed an empirical equation $C = C_I \exp(-k\theta)$.

Wen and Hashinger⁵⁾ measured a system similar to that of Leva and correlated the data by dimensionless terms concerning the fluidizing characteristics. An empirical equation has been derived within $\pm 100\%$ deviation.

Yagi and Kunii⁶⁾ have correlated their data for continuous and batch fluidized beds within $\pm 50\%$ deviation.

In our previous paper³⁾, the effects of apparatus structure (type of column top, type of overflow pipe, height of feed pipe from distributor and type of distributor) on the modified elutriation rate coefficients were measured.

In this work, the effects of fluid properties (air and water), superficial velocity, feed rate of solids, fines

concentration in feed, particle size distribution and so on, on the elutriation rate of fines from continuous and batch fluidized beds were examined.

In particular, it was found from experiments in a continuous fluidized bed of two particle-size system that the concentration of fines at is kept a constant value C_s throughout the bed height except in the space near the overflow pipe, and that C_s is generally smaller than the fines concentration in the overflow pipe C_0 .

An empirical dimensionless equation which correlates the elutriation rate coefficients on the basis of C_s has been derived within $\pm 25\%$ deviation.

1. Experimental Apparatus and Method

1-1 Continuous and batch fluidized beds of gas-solid system

Experimental apparatus is shown in **Fig. 1**. The apparatus for the measurement of the axial distribution of fines concentration is shown in **Fig. 2**.

For batch operations, 8, 10 and 11 in **Fig. 1** were not used.

The air led from the blower passed through the surge tank, the quadrant nozzle (for measuring flow rate) and the control valve into the fixed bed (glass beads and lead balls of 1~3 mm in diameter were fixed up to 25 cm height) and went through the distributor† (perforated plate with hole diameter of 1.5 mm in 3 mm pitch, covered with a 200 mesh screen) into the fluidized bed.

On the other hand, in the continuous fluidized bed, the fluidized particles arranged for a desired fines concentration were fed to the hopper and brought to the fluidized bed by the screw feeder which was regulated by a variable-speed motor. In the batch system the particles were fed to the fluidized column from the top.

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† No effect of the kind of distributor on the elutriation rate has been confirmed in the previous work.³⁾

Table 1 Fluidized particles

No.	Particles	System	Range [mesh]	Density ρ_s [g/cm ³]	Terminal velocity u_t [cm/sec]	Diameter* D_p [μ]
1	Glass beads	Liquid-Solid	100—150	2.55	0.805	106
	Soma sand		22—28	2.67	11.7	718
2	Toyoura sand		70—80	2.67	2.14	178
	Soma sand		22—28	2.67	11.7	718
3	Stainless balls		30—50	7.47	20.4	505
	Lead balls		8—12	11.1	77.7	1930
4	Glass beads	Gas-Solid	100—150	2.50	87	135
	Soma sand		22—28	2.67	588	718
5	Toyoura sand		70—80	2.67	127	178
	Soma sand		22—28	2.67	588	718
6	Neobeads**		80—100	1.54	64	141
	Soma sand		22—28	2.67	588	718
7	Neobeads**		50—250	1.54	64	140
8	Toyoura sand		50—250	2.67	118	165

* D_p = median diameter. Refer to the footnote*

** Neobeads are porous and spherical particles.

The weight of particles was divided by their apparent volume occupied to obtain ρ_s .

Fig. 1 Schematic diagram of the continuous fluidized bed of gas-solid system

1-2 Batch fluidized bed of liquid-solid system

The experimental apparatus is shown in Fig. 3. The water led from the constant level head tank passed through the control valve, the fixed bed (glass beads of about 5 mm in diameter were fixed) and the distributor (which was the same plate as mentioned in 1-1 and was set on the glass beads fixed bed) into the fluidized bed.

The water containing the fines elutriated was led to the solid separator from the top of the column and discharged after measurement of the water flow rate.

* The particle diameter D_p was measured by the air elutriation method for the fine particles of Nos. 1, 2, 4, 5 and 6 and by the water elutriation method for the fines of No. 3 and the coarse particles of No. 1~No. 6.

The terminal velocity u_t was defined as fluid velocity u for the situation where half of the initial weight of particles was elutriated and D_p was calculated from this u_t by the use of Re number vs. drag coefficient table.

D_p and u_t of Nos. 7 and 8 were calculated by the method mentioned later in 4-2-2.

Fig. 2 Measurement apparatus of fines concentration

Various column heights and diameters were used to investigate their effects on the elutriation rate coefficients.

The fluidized particles used are shown in Table 1*.

2. Definition of Various Elutriation Rate Coefficients

2-1 Two particle-size system

2-1-1 Batch fluidized bed Leva¹⁾ has proposed the following equation for elutriation of fines,

$$dC/d\theta = -kC \quad (1)$$

where C represents the fines concentration in the bed.

k is the elutriation rate coefficient. It varies proportionally with the cross-sectional area of the bed and inversely with the bed height. Wen *et al.*⁵⁾

Fig. 3 Schematic diagram of the batch fluidized bed of liquid-solid system

and Yagi *et al.*⁶⁾, therefore, have defined the modified elutriation rate coefficient K^* as follows.

$$k = K^* A / W \quad (2)$$

From Eqs.(1) and (2)

$$dC/d\theta = -K^* AC / W \quad (3)$$

From the total mass balance,

$$V = -dW/d\theta \quad (4)$$

and from the mass balance of fines,

$$V = -d(WC)/d\theta \quad (5)$$

From Eqs.(3)~(5),

$$V = K^* AC / (1 - C) \quad (6)$$

2-1-2 Continuous fluidized bed By replacing C with C_0 in Eq.(6), K_0^* is defined by

$$V = K_0^* AC_0 / (1 - C_0) \quad (7)$$

Similarly, by replacing C with C_s , K_s^* is defined by

$$V = K_s^* AC_s / (1 - C_s) \quad (8)$$

Yagi *et al.*⁶⁾ have proposed that V is proportional to C_s and that $\bar{K}_s^*$ is defined by

$$V = \bar{K}_s^* AC_s \quad (9)$$

2-2 Multi particle-size system

2-2-1 Batch fluidized bed Applying Eq.(3) to a multi particle-size system, one gets

$$\frac{d\left(\int_0^x K_F(x) dx\right)}{d\theta} = -\frac{K^* A \left(\int_0^x K_F(x) dx\right)}{W} \quad (10)$$

When the particle diameter is less than x , the mass balance is written as

$$V_i = V \left(\int_0^x K_E(x) dx\right) = -d \left(W \int_0^x K_F(x) dx \right) / d\theta \quad (11)$$

where $\int_0^x K_E(x) dx$ is the fines concentration in the elutriated solids and $\int_0^x K_F(x) dx$ is the fines concentration in fluidized bed.

From Eqs.(4), (10) and (11), one gets

* They used $C_s \div C_0$ in the calculation of K_s^* .

$$V_i = \frac{K^* A \left(\int_0^x K_F(x) dx\right)}{\left(1 - \int_0^x K_F(x) dx / \int_0^x K_E(x) dx\right)} \quad (12)$$

2-2-2 Continuous fluidized bed By replacing K^* with K_s^* in Eq.(12), K_s^* is defined by

$$V_i = \frac{K_s^* A \left(\int_0^x K_F(x) dx\right)}{\left(1 - \int_0^x K_F(x) dx / \int_0^x K_E(x) dx\right)} \quad (13)$$

Applying Eqs.(12) and (13) to the batch and continuous fluidized beds of two particle-size system with Eqs.(14) and (15),

$$\int_0^x K_E(x) dx = 1 \quad (14)$$

$$\int_0^x K_F(x) dx = \begin{cases} C & \text{(batch)} \\ C_s & \text{(continuous)} \end{cases} \quad (15)$$

one gets $V_i = V = K^* AC / (1 - C)$ (batch)

and $V_i = V = K_s^* AC_s / (1 - C_s)$ (continuous)

which are identical with Eq.(6) and Eq.(8), respectively.

3. Dimensional Analysis of Elutriation Rate

Following three equations may be considered as the basic equations for elutriation.

For the momentum balance of single particle

$$-\rho_s \frac{\pi D_p^3}{6} \frac{\partial u_s}{\partial \theta} = \frac{\pi D_p^3 (\rho_s - \rho) g}{6} - C_D \frac{\pi D_p^2}{4} \frac{(u_s - u_F)^2 \rho}{2} \quad (16)$$

For the momentum balance of fluid in column

$$\rho \frac{\partial u_F}{\partial \theta} = -\frac{\partial p}{\partial z} + \mu \frac{1}{r} \frac{\partial}{\partial r} \left(r \frac{\partial u_F}{\partial r} \right) + \rho g \quad (17)$$

For elutriation

$$V = -dW/d\theta = K^* AC / (1 - C) \quad (18)$$

The following dimensionless quantities are introduced:

$$\begin{aligned} u_s^* &= u_s / (u - u_t), \quad \theta^* = \theta (u - u_t) / D_p, \quad u^* = u_F / (u - u_t) \\ W^* &= W / \rho D_p^3, \quad p^* = (p - p_0) / \rho (u - u_t)^2, \quad r^* = r / D_p, \\ Z^* &= Z / D_p \end{aligned} \quad (19)$$

Then, the above three equations become

$$-\frac{\partial u_s^*}{\partial \theta^*} = \left(\frac{\rho_s - \rho}{\rho_s} \right) \frac{g D_p}{(u - u_t)^2} - C_D \frac{3}{4} \left(\frac{\rho}{\rho_s} \right) (u_s^* - u^*)^2 \quad (20)$$

$$\begin{aligned} \frac{\partial u^*}{\partial \theta^*} &= -\frac{\partial p^*}{\partial z^*} + \frac{\mu}{D_p (u - u_t) \rho} \frac{1}{r^*} \frac{\partial}{\partial r^*} \left(r^* \frac{\partial u^*}{\partial r^*} \right) \\ &\quad + \frac{g D_p}{(u - u_t)^2} \end{aligned} \quad (21)$$

$$-\frac{dW^*}{d\theta^*} = \frac{K^*}{\rho (u - u_t)} \frac{\pi}{4} \left(\frac{D}{D_p} \right)^2 \frac{C}{(1 - C)} \quad (22)$$

From Eqs.(20)~(22), one gets

$$\begin{aligned} \frac{K^*}{\rho (u - u_t)} &= f \left(\frac{\rho_s - \rho}{\rho_s}, \frac{g D_p}{(u - u_t)^2}, C_D, \frac{\rho}{\rho_s}, \right. \\ &\quad \left. \frac{D_p (u - u_t) \rho}{\mu}, \frac{D}{D_p} \right) \end{aligned} \quad (23)$$

Since $(\rho_s - \rho) / \rho_s$ and ρ / ρ_s are not independent, only $(\rho_s - \rho) / \rho$ is used. As C_D is a function of $D_p (u - u_t) \rho / \mu$, C_D is eliminated in Eq.(23).

Finally, the following equation is obtained:

($u - u_t = 30 \text{ cm/sec}$, $C_f = 0.5$, Particles No. 5)

Fig. 4 Longitudinal distribution of fines concentration

($C_f = 0.2$, Particles No. 5)

Fig. 5 Relationship between C_s/C_0 and F

$$\frac{K^*}{\rho(u - u_t)} = f\left(\frac{\rho_s - \rho}{\rho}, \frac{(u - u_t)^2}{gD_p}, \frac{D_p(u - u_t)\rho}{\mu}, \frac{D}{D_p}\right) \quad (24)$$

4. Experimental Results and Discussion

4-1 Axial distribution of fines concentration in gas-solid continuous fluidized bed of two particle-size system

As shown in Fig. 2, the sampling pipes were designed to take samples of solids from the bed at height intervals of 10 mm.

A typical example of the axial distribution of fines concentration is shown in Fig. 4. Tendencies similar to those in Fig. 4 were obtained in every C_f .

Fig. 4 shows that fines concentration is constant throughout the bed height except in the space near the overflow pipe, where it is greater than the constant value within the bed.

The radial distribution of fines concentration was found to be uniform by inserting the movable sampling pipe into the bed through the fixed sampling pipe in Fig. 2.

Relationship between C_s/C_0 and feed rate F is shown

(Particles No. 5)

Fig. 6 Effects of feed rate on the modified elutriation rate coefficients, K_s^* , $\bar{K}_s^*$ and K_0^*

Fig. 7 Relationship between K^* , K_s^* and $u - u_t$

in Fig. 5. As F becomes larger, C_s/C_0 approaches unity.

4-2 Modified elutriation rate coefficients

4-2-1 Gas-solid fluidized bed of two-particle size system The values of K_0^* , K_s^* and $\bar{K}_s^*$ calculated from Eq.(7), Eq.(8) and Eq.(9) respectively using C_0 and C_s obtained in 4-1 are shown in Fig. 6.

K_s^* is not affected by feed rate F at any C_f . The effect of F on $\bar{K}_s^*$ does not appear at $C_f = 0.2$ and 0.5 , but it does at $C_f = 0.7$ when $u - u_t$ is small.

Fig. 8-a Relationship between K^* and $u - u_t$

Fig. 8-b Relationship between K^* and $u - u_t$

Fig. 9 $K^*/\rho(u - u_t)$ or $K_s^*/\rho(u - u_t)$ vs. Fr

Fig. 10 $K^*/\rho(u - u_t)Fr^{0.5}$ or $K_s^*/\rho(u - u_t)Fr^{0.5}$ vs. Re

As described in the previous report³⁾, K_0^* is affected by the structure of the overflow pipe and moreover by F at every C_f , as shown in Fig. 6. Therefore, a modified elutriation rate coefficient must be calculated based on C_s instead of C_0 and Eq.(8) is better than Eq.(9) because K_s^* is not so much affected by F as $\bar{K}_s^*$. Moreover, Fig. 6 shows that C_f has no effect on K_s^* , $\bar{K}_s^*$ and K_0^* when $u - u_t$ is large.

4-2-2 Gas-solid fluidized bed of multi particle-size system The size distribution function for the particles in the overflow pipe, generally, seems to be different from that in the fluidized bed.

In this experiment, however, both distribution functions were found to be almost the same, because the experiment was performed under conditions where $C_0 \doteq C_s$ (refer to Fig. 5).

In these cases, the value of 50% pass diameter in the particle-size cumulative distribution curve from the minimum diameter to a certain diameter x , was used as representative particle diameter D_p , and u_t was calculated by the use of this value of D_p . Fig. 7 shows these results.

It is found from Fig. 7 that K^* and K_s^* can be correlated by the same method.

4-2-3 Liquid-solid batch fluidized bed of two particle-size system Relationship between K^* and $u - u_t$

is shown in Figs. 8-a and b.

It is recognized from Figs. 8-a and b that the experimental data of K^* deviate upward when the column is shorter than a certain height. It seems to be affected mainly by fines that are carried out with fluidizing fluid in the case of a shorter column and that are going back to the bed by gravity in the case of a longer column.

Fig. 8-b, moreover, shows no effect of column diameter on K^* as in the previous report ($D=67, 150$ mm).

It is also found that T.D.H. (Transport Disengaging Height), namely, the height at which K^* is not affected by column height exists.

4-3 An empirical dimensionless equation

When the data were correlated, it was found that $\frac{D_p u_t \rho}{\mu}$ was better than $\frac{D_p (u - u_t) \rho}{\mu}$ in Eq.(24), and the former is used as Reynolds number. D/D_p was eliminated because D has no effect on K^* in the present or the previous works³⁾.

Dependences of $K^*/\rho(u - u_t)$ or $K_s^*/\rho(u - u_t)$ on Froude number, Reynolds number and $(\rho_s - \rho)/\rho$ are shown in Figs. 9, 10 and 11.

Fig. 12 shows all data in this work and data of batch fluidized bed in other works. Some data of the latter cases have larger deviations than $\pm 25\%$. But the results of Leva¹⁾ originally have large scattering. Other investigators' data^{5,7)} are recalculated from figures (similar to Fig. 9) in the original papers. These also give slightly larger deviation than $\pm 25\%$.

Data of Wen *et al.*⁵⁾ have an upward deviation from

Fig. 11 $K^*/\rho(u-u_t)Fr^{0.5}Re^{0.3}$ or $K_s^*/\rho(u-u_t)Fr^{0.5}Re^{0.3}$ vs. $(\rho_s - \rho)/\rho$

Fig. 12 Correlation for modified elutriation rate coefficients K^* and K_s^* in gas-solid, liquid-solid systems

those of this study. This may be caused by their use of the height below T.D.H. in their fluidized bed. Other reports on the continuous fluidized bed cannot be directly compared with this work because of the difference of correlating method.

Conclusion

(1) The axial distribution of fines concentration was measured in a continuous gas-solid fluidized bed with two particle-size system. It was found that the fines concentration is kept at a constant value C_s throughout the bed height except in the space near the overflow pipe, and that C_s is smaller than the fines concentration in the overflow pipe C_0 .
(2) The modified elutriation rate coefficients K_0^* , K_s^* and $\bar{K}_s^*$ defined by Eq.(7), Eq.(8) and Eq.(9) respectively were examined. K_0^* was affected by the structure of the overflow pipe and feed rate F , and $\bar{K}_s^*$ was also affected by F when the fines concentration in feed C_f was large and $u-u_t$ was small. But K_s^* defined by Eq.(8) was not affected by F at any C_f . Therefore, K_s^* is better than K_0^* or $\bar{K}_s^*$. K^* calculated by Eqs.(1) and (2) in a batch fluidized bed can be correlated similarly to K_s^* .

(3) It was found that Transport Disengaging Height (T.D.H.) exists.

(4) K^* and K_s^* obtained with the column height larger than T.D.H. for the cases with two particle-size and multi particle-size, liquid-solid and gas-solid systems, and batch and continuous operations, have been correlated within $\pm 25\%$ deviation, as shown in Fig. 12.

(5) It was found in the present and the previous work³⁾ that column diameter ($D=31, 67, 150$ mm) has no effect on K^* and K_s^* .

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Nomenclature

A	= cross-sectional area of fluidized bed	[cm ²]
C	= fines concentration in fluidized bed	[—]
C_D	= friction factor	[—]
C_f	= fines concentration in feed	[—]
C_I	= initial fines concentration in fluidized bed	[—]
C_0	= fines concentration in overflow	[—]
C_s	= constant fines concentration in fluidized bed	[—]
D_p	= particle diameter	[cm]
D	= bed diameter	[cm]
F	= feed rate of solids	[g/sec]
g	= gravitational constant	[cm/sec ²]
H	= height of bed from distributor	[cm]
k	= elutriation rate coefficient	[1/sec]
K^*	= modified elutriation rate coefficient for batch fluidized bed	[g/cm ² ·sec]
K_0^*	= modified elutriation rate coefficient in Eq.(7)	[g/cm ² ·sec]
K_s^*	= modified elutriation rate coefficient in Eq.(8)	[g/cm ² ·sec]
$\bar{K}_s^*$	= modified elutriation rate coefficient in Eq.(9)	[g/cm ² ·sec]
$K_E(x)$	= size distribution function for elutriated particles	[1/cm]
$K_F(x)$	= size distribution function for particles in fluidized bed	[1/cm]
p	= pressure	[g/cm·sec ²]
r	= radial distance	[cm]
u	= superficial fluid (air or water) velocity	[cm/sec]
u_F	= fluid (air or water) velocity	[cm/sec]
u_s	= particle velocity	[cm/sec]
u_t	= terminal velocity of solid particle	[cm/sec]
V	= elutriation rate	[g/sec]
V_i	= elutriation rate of particle of which diameter is less than x	[g/sec]
W	= holdup	[g]
x	= particle diameter	[cm]
Z	= height of column from distributor	[cm]
ρ	= fluid density	[g/cm ³]
ρ_s	= particle density	[g/cm ³]
θ	= time	[sec]
μ	= fluid viscosity	[g/cm·sec]
Re	= $D_p u_t \rho / \mu$	[—]
Fr	= $(u-u_t)^2 / g D_p$	[—]

Literature Cited

- 1) Leva, M.: *Chem. Eng. Progr.*, **47**, 39 (1951)
- 2) Osberg, G. and D. H. Charlesworth: *Chem. Eng. Progr.*, **47**, 566 (1951)
- 3) Tanaka, I., H. Shinohara, H. Hirose and Y. Tanaka: *Technology Reports of the Kyushu University*, **44**, No. 3, 406 (1971)
- 4) Nagamatsu, S., Y. Tsuru, Y. Tanaka, H. Shinohara and M. Wada: *Kagaku Kōgaku (Chem. Eng., Japan)*, **25**,

- 2 (1961)
 5) Wen, C. Y. and R. F. Hashinger: *A.I.Ch.E. Journal*,
 6, 220 (1960)
 6) Yagi, S. and D. Kunii: *Chem. Eng. Sci.*, 16, 364, 372,

- 380 (1961)
 7) Yagi, S. and T. Aochi: paper presented at the Spring
 Meeting of the Society of Chemical Engineers, Japan
 (1955)

ELUTRIATION OF FINES FROM FLUIDIZED BED*

— Study of Transport Disengaging Height —

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Transport Disengaging Height (T.D.H.) was studied theoretically and experimentally for both continuous and batch fluidized beds.

The theoretical elutriation rate of fines above the T.D.H., V_t , was calculated by use of the Maxwell-Boltzmann energy distribution, and it showed good agreement with the experimental value for the continuous fluidized bed.

The theoretical values of the T.D.H., which were defined as Z at $V=1.01V_t$, coincided fairly well with the present and reported experimental results.

The elutriation height coefficients a were found empirically to increase with increase in particle diameter and with decrease in superficial velocity.

Introduction

It is important to study elutriation of fine particles from fluidized beds used for reaction, drying and classifying.

In our previous work⁵⁾, the elutriation rate above the T.D.H., V_t , that is, the elutriation rate at a column height larger than the T.D.H., was studied experimentally and the modified elutriation rate coefficients K_s^* and K^* obtained in gas-solid and liquid-solid systems and in batch and continuous fluidized beds were correlated within $\pm 25\%$ deviation by a dimensionless equation.

In this work, the T.D.H., an important factor in the design of fluidized beds, was studied experimentally and analytically for continuous and batch fluidized beds. For the maintenance of steady state, the latter was operated with total reflux of elutriated fines.

There are several previous investigations. Zenz and Weil⁸⁾ have shown how to determine the T.D.H. from the figure obtained from an experimental study.

Andrews¹⁾ has derived an analytical relationship between the elutriation rate V and the height from the dense bed surface Z by use of the Maxwell-Boltzmann energy distribution as follows:

$$-\partial V / \partial Z = \lambda A \beta \rho \sqrt{2g} mg \exp(-\beta mg Z) \quad (1)$$

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where $\lambda A \beta \rho \sqrt{2g} mg$ and βmg must be determined experimentally. The relationship between these factors and the particle characteristics or the operating conditions was not given.

Blyakher and Pavlov²⁾ studied the dependence of the T.D.H. on grids located above a conical fluidized bed.

Lewis *et al.*⁴⁾ have proposed an empirical equation as follows:

$$V = K \exp(-aZ), \quad K = cuA \exp[-(b/u)^2] \quad (2)$$

where a , b and c are the experimental coefficients, but the reported values of c are scattered. The above coefficient a is called the elutriation height coefficient especially by the authors.

Kunii and Levenspiel³⁾ have derived an analytical equation as follows

$$V/V_0 = 1 - (1 - V_s/V_0)[1 - \exp(-aZ)] \quad (3)$$

The elutriation rate above the T.D.H. which they used was the saturation carrying flow rate V_s that was the maximum elutriation rate observed under pneumatic transport conditions of the single particle-size system by Zenz *et al.*⁸⁾. However, it is found that the elutriation rate above the T.D.H., V_t , in this study is much smaller than V_s .

They gave no experimental values of a and V_0 .

In this work, the elutriation rate above the T.D.H., V_t , has been analysed by use of the Maxwell-Boltzmann energy distribution. This V_t should be used in place of V_s in Eq.(3). The elutriation height coefficient a proposed by Lewis *et al.*⁴⁾ and Kunii *et al.*³⁾ was measured experimentally and an equation giving the value of the T.D.H. has been derived analytically.