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ELUTRIATION OF MIXED FEED IN A TOP FEED CONTINUOUS FLUIDIZED BED

by

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TO

MY MOTHER

ABSTRACT

Elutriation data comparing batch and continuous systems is reported for coal char powder. The results were obtained for a 2.5 cm bed diameter with continuous feed from the top and product removal from the bottom. The coal char feed mixture used for the studies contained nearly 80 percent of particles ranging between 300 and 45 microns. The remaining 20 percent were fines having a particle diameter less than 45 microns.

Elutriation rate constants have been calculated for three particle sizes of (a) less than 45 microns, (b) 45 to 53 microns and (c) 53 to 75 microns. The dependence of elutriation rate constant on process variables such as air velocity and bed height is discussed. It was observed that

$$\text{elutriation rate constant} \propto (\text{air velocity})^n$$

and the factor n was found to be dependent on bed height. The values

of n obtained for particle size of 75 to 53 microns were 3.33, 2.78 and 1.86 for bed heights of 8.5, 17.5 and 22.0 cm respectively. A similar order of values were obtained for other particle sizes. In the literature, for batch systems, the value of the factor n has been reported to be between 2.0 and 4.0 and independent of bed height. In a few batch runs, conducted in this work, for 22.0 cm bed height the factor n was calculated to be 3.32, 3.44 and 3.34 for particle sizes of 75 to 53 microns, 53 to 45 microns and less than 45 microns respectively. A bed height of 22.0 cm was chosen for batch runs because for continuous system, the values of n were lowest for this bed height.

Results obtained for the continuous system have been compared with batch data obtained together with batch data reported in the literature. A theoretical basis has been suggested for the differences observed in continuous and batch elutriation schemes, based on a bubble model for countercurrent backmixing.

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NOMENCLATURE

A	Bed crossectional area, cm^2
D_p	Particle diameter, microns (μ)
G	Mass flow rate, gm/sec ~~cm^2~~
g	Acceleration due to gravity, 981 cm/sec^2
g_c	Conversion factor, 981 gm cm/gmf sec^2
k	Elutriation rate constant, $time^{-1}$
$\bar{t}$	Residence time, min
u	Velocity, cm/sec
W	Bed weight, gm
ρ	Density, gm/cc
μ	Viscosity of gas, gm/cm sec

Subscripts

s	solids
g	gas
i	particle size
t	terminal value

I. INTRODUCTION

Elutriation is a process in which smaller particles are continuously removed by a fluid stream passing through a fluidized bed. In fluidized process technology, the size of particles is of extreme importance. Raw materials generally available consist of a range of particle sizes. Hence, removal of the undesired size particles becomes an important step in any fluidized bed operation. In this particular step, knowledge of the elutriation behaviour helps in design.

In the past, several studies(1) have been conducted to predict the behaviour of elutriated particles in batch systems. However, it is felt that batch system data may not apply to a continuous elutriation scheme. In a batch system, the fluidized bed keeps on depleting in fines while in the continuous system, a steady state composition is reached. The present work was done to observe this difference for an air coal-char system.

2. LITERATURE SURVEY

A literature survey conducted for continuous elutriation showed that very little information is available for large diameter beds. Tanaka and Yoshimura (1962)[2] have reported a study on bed diameter of 16. 7 cm. They emphasise the effect of location of feed and product outlet points on elutriation rate. Sufficient data is available on batch elutriation systems. An excellent review is presented by Davidson and Harrison (1971)[1] in the chapter on elutriation.

2.1 Mechanism of Elutriation

To understand the differences between continuous and batch elutriation, it is necessary to have a clear picture of bed behaviour. A summary of the mechanism proposed by Rowe and Partridge (1965)[3] is presented here which will be used later to explain the different behaviour of continuous system.

Gas bubbles formed just above the distributor plate rise up carrying solid particles in the wake as shown in figure 1. As these bubbles travel upwards, solids are being continually exchanged with fresh solids in the dispersed phase. The velocity of bubble rise is considerably larger than the superficial gas velocity. As these bubbles reach the bed surface, sudden eruptions occur and solid particles are thrown in the freeboard. Air velocity in the freeboard is considerably less than the bubble velocity so larger particles return to the bed while smaller ones having terminal velocity less than the superficial air velocity are removed alongwith the gas stream. At the bed surface, the velocity profile of air is not uniform due to eruption of bubbles. After a certain height, known as the transport disengagement height (TDH), the velocity profile assumes its parabolic distribution. Hence, for any elutriation apparatus, the freeboard must be larger than TDH to study the effects of process variables.

By the above mechanism, it follows that mixing patterns in the bed influence the elutriation rate significantly. For a larger degree of stratification (orientation of particles with respect to size in a vertical plane) the elutriation rate should be higher, as bubbles erupting at the surface will have a greater number of smaller particles. For prediction of mixing behaviour in the fluidized bed, Kunii and Levenspiel (1968)[4]

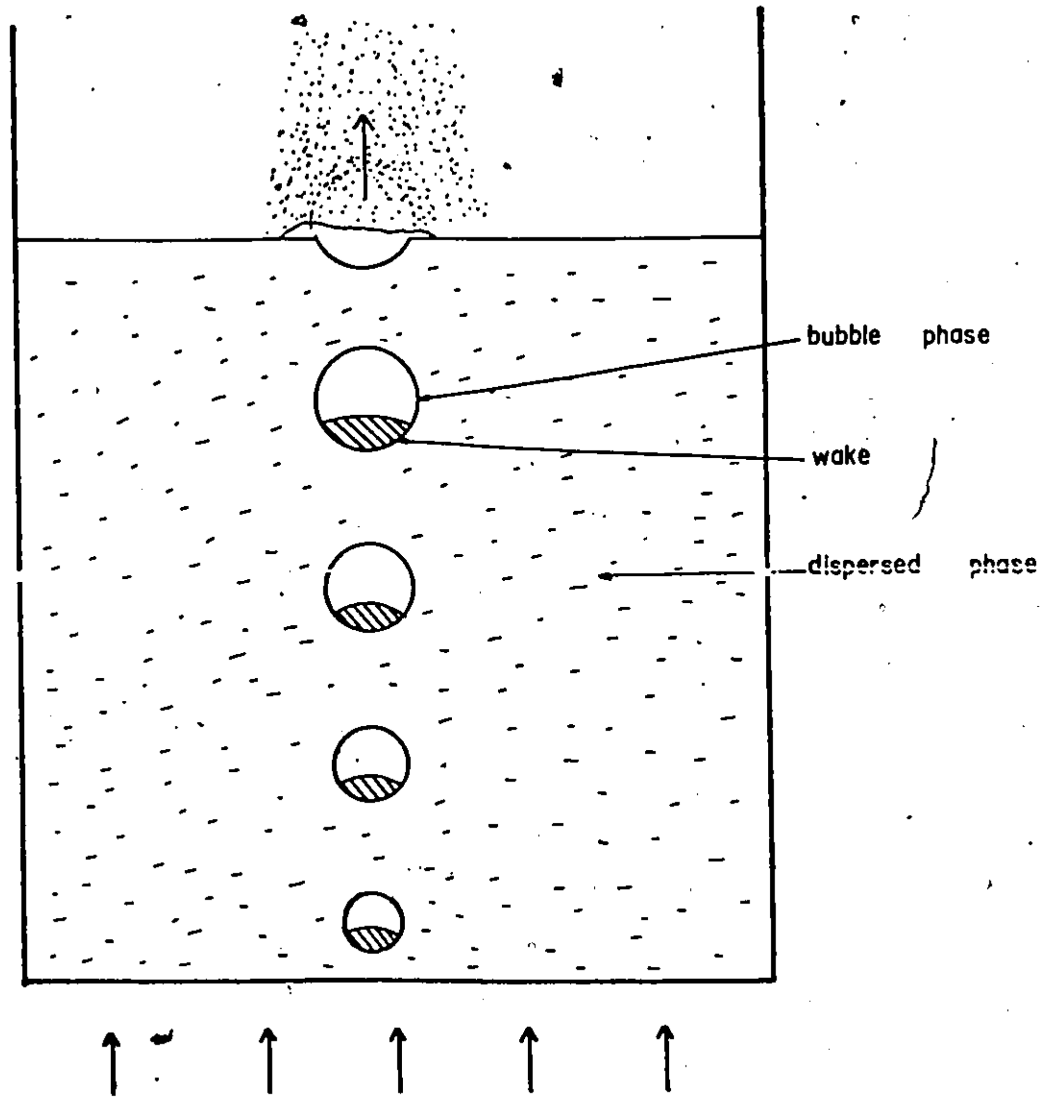


Fig. 1

Schematic diagram of particle elutriation due to gas bubbles(3) have proposed a countercurrent backmixing model. Qualitatively, it states that movement of bubbles displaces solids upwards leading to a downward movement of solids in the remainder of the bed. This downward movement of the dense phase causes a counter current flow of gas and solids. As the solids travel downwards, exchange takes place with the material in the bubble wake. Thus, finer particles tend to accumulate at the top while larger ones sink to the bottom. However, in some cases, the downward solids flow may be enough to pull some gas bubbles down and thus provide backmixing for gas phase also.

Osberg and Charlesworth (1951)[5] reported some batch stratification data as shown in figure 2, which shows that there is a sharp concentration gradient in the top 1 cm of the bed and most of the fines are accumulated in that region. The effect of stirrers in the bed was reported by Lewis et al. (1962)[6] but no convincing explanation was provided for the decrease of elutriation rate with stirring. Figure 3 provides a reproduction of their results. Elutriation rate per unit superficial gas velocity is plotted against the square of the inverse of superficial gas velocity on semi logarithmic plot. The results clearly show a decrease in elutriation rate with stirring. Wen and Hashinger (1960)[7] studied the effects of horizontal baffles, made of wire mesh, in a fluidized bed of 2 in. diameter. Figure 4 shows their results. The presence of baffles does not affect the elutriation rate constant (k) for lower bed

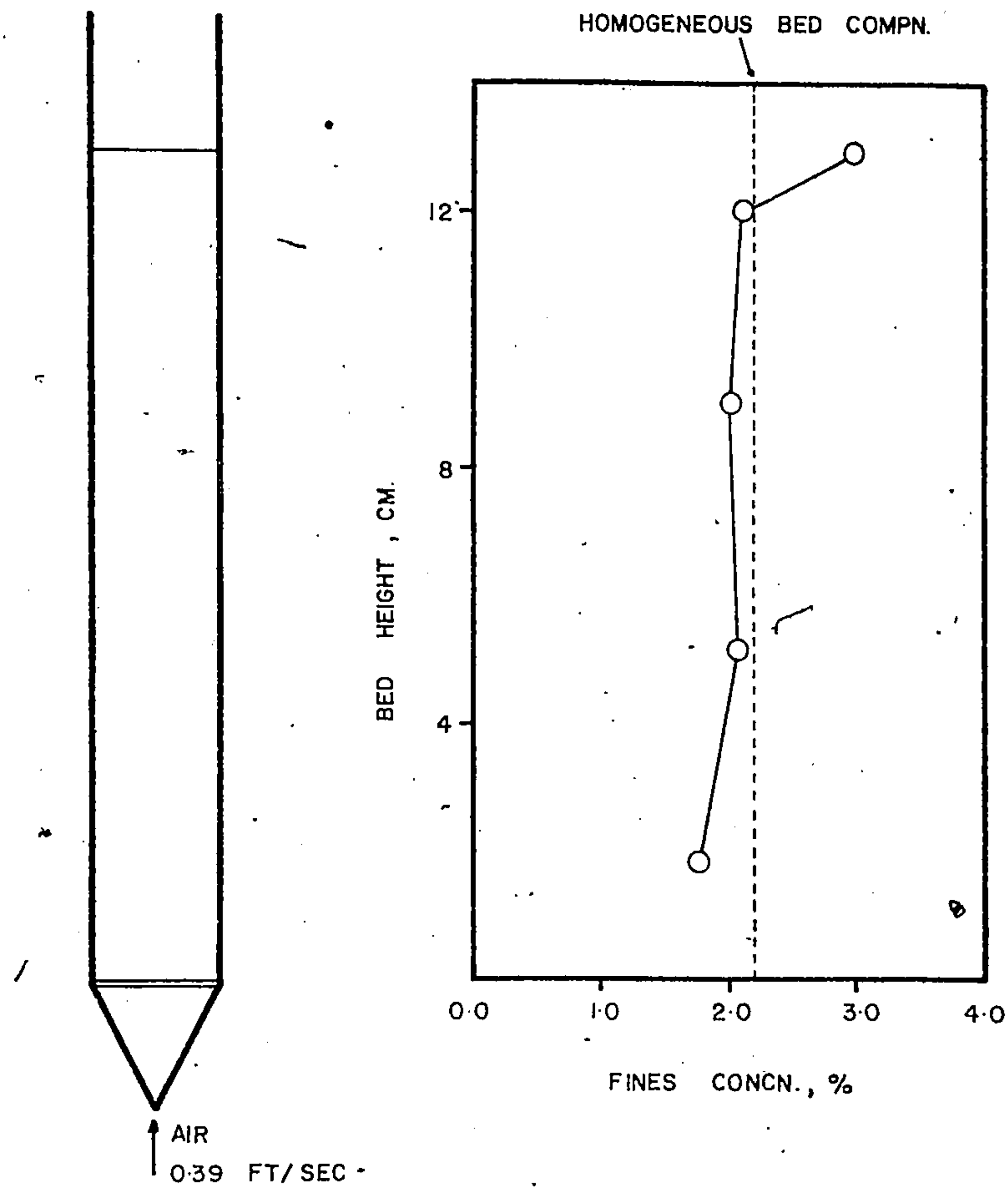


Fig. 2

Bed stratification data for mixture containing (-150 +200) mesh[5)]

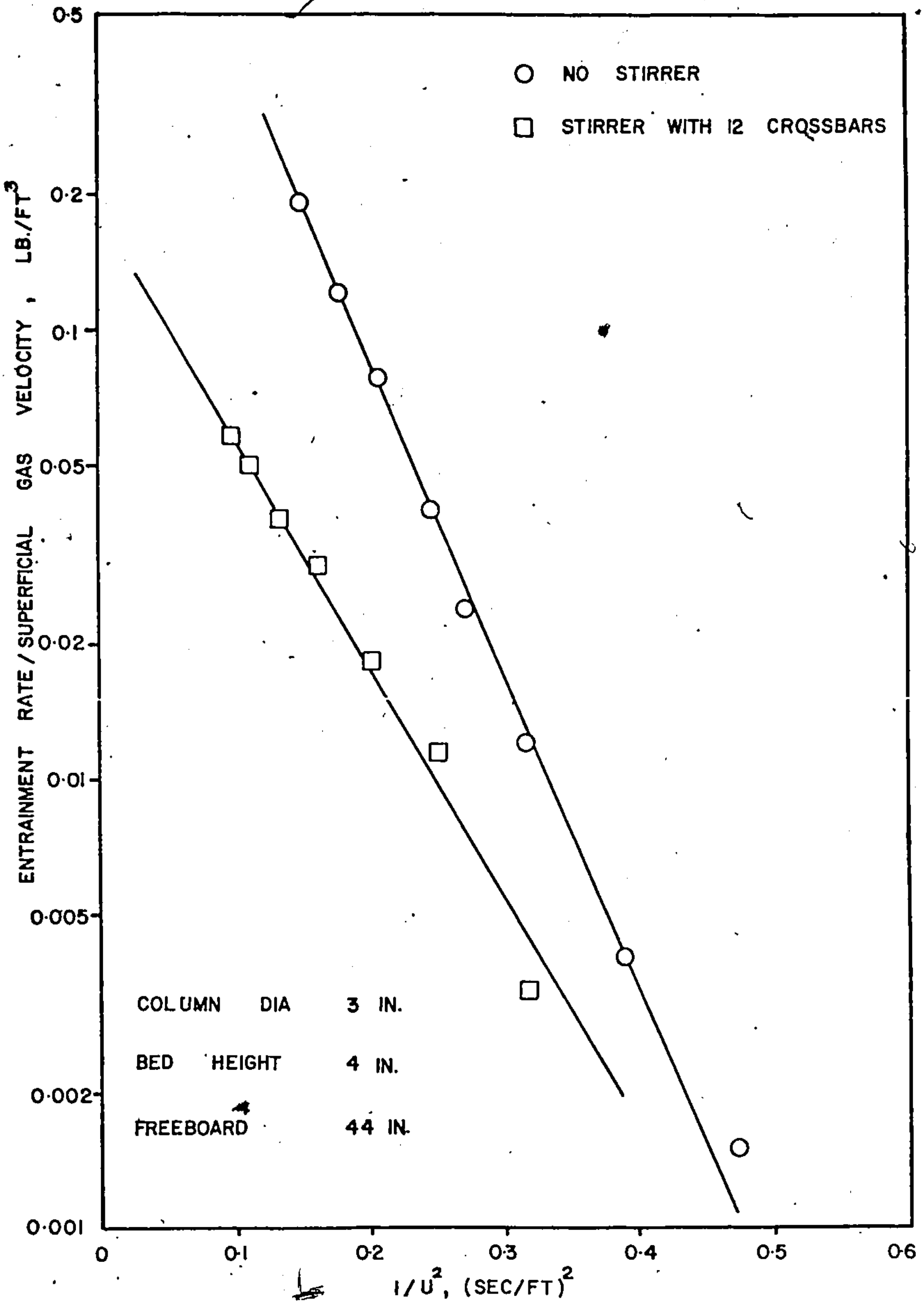


Fig. 3 - Effect of stirrer on entrainment of 75μ glass spheres[6]

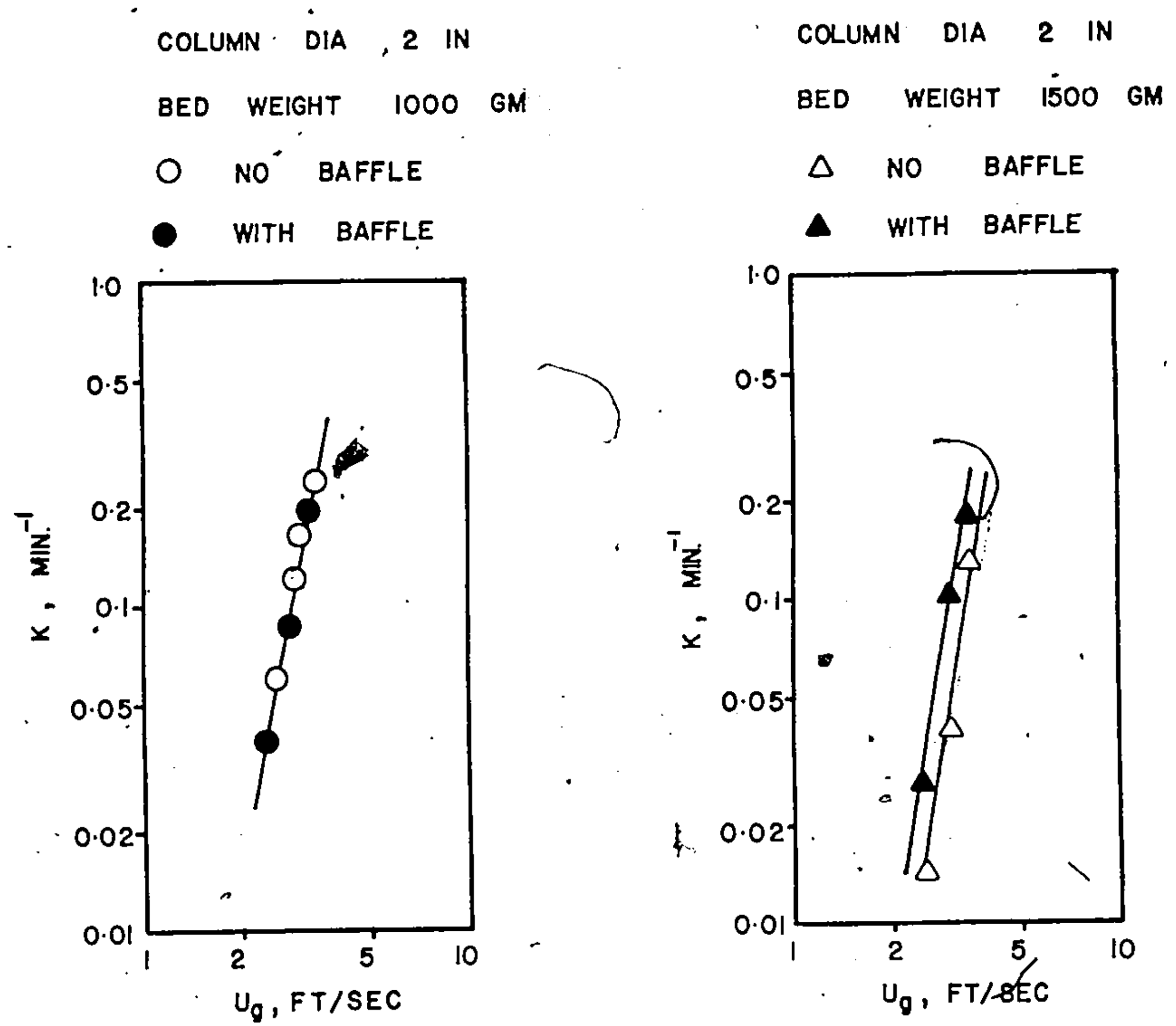


Fig. 4

Effect of a baffle on the elutriation rate constant(7) heights. However, for larger bed heights, k is increased. A theoretical explanation for these effects is discussed in the results and discussion section with reference to the results obtained for continuous system.

Wen and Hashinger (1960)[7] have also reported the effects of column diameter and freeboard on elutriation rate. They conclude that elutriation rate is independent of these parameters if freeboard is more than the TDH and column diameter is greater than 10 cm. The TDH as a function of superficial air velocity and bed diameter has been reported by Zenz and Othmer (1960)[8] but the data reported is too limited to have any practical use. Only an approximate value of TDH can be estimated by extrapolation of their reported curves.

2.2 Mathematical Analysis of Elutriation

Levenspiel (1962)[9] proposed some relations to predict elutriation behaviour quantitatively. These equations have been recommended for the design of a fluidized bed reactor. He proposed that the amount of particles of a given size blown out from the bed shows a first order behaviour with respect to bed concentration. Mathematically,

$$\frac{d}{dt}\,(\text{number of marked particles}) = k(\text{number of marked particles}) \tag{i}$$

where k is termed as the elutriation rate constant. k is dependent on system parameters such as gas velocity, bed height and particle size. An approximate correlation suggested by Levenspiel (1962)[9] is

$$k \propto \frac{(\text{gas velocity})^4}{(\text{bed height})\,(\text{particle size})^{2 \text{ to } 3}} \qquad \text{(ii)}$$

Wen and Hashinger (1960)[7] have reported an empirical correlation for batch systems.

$$\frac{kA}{W\rho_g(u_g - u_t)} = 1.52 \times 10^{-5} \left[\frac{(u_g - u_t)^2}{g D_p}\right]^{0.5} \left[\frac{D_p u_t \rho_g}{\mu}\right]^{0.725} \left[\frac{\rho_s - \rho_g}{\rho_g}\right]^{1.15} \qquad \text{(iii)}$$

The limits of operation for the above correlation are

$$0.001595 \text{ in} < D_p < 0.0058 \text{ in}$$

$$0.0103 \text{ lb/cu ft} < \rho_g < 0.075 \text{ lb/cu ft}$$

$$81 \text{ lb/cu ft} < \rho_s < 312 \text{ lb/cu ft}$$

$$0.725 \text{ ft/sec} < u_g < 4.33 \text{ ft/sec}$$

A description of the various symbols used in equation (iii) is provided in the nomenclature. Equation (iii) when reduced for gas velocity dependence of k gives

$$k \propto (u_g - u_t)^2 \tag{iv}$$

From equation (ii) and (iv) it is observed that air velocity is the most important process variable affecting the elutriation rate constant and hence, the elutriation rate.

The following relations are used to calculate the constant k and the residence time of particle of size i in an experimental run.

$$k_i = \frac{\text{rate of carryover of particle size } i}{\text{wt. of such particles present in bed}} \tag{v}$$

$$\bar{t}_i = \frac{\text{wt. of particles of size } i \text{ in bed}}{\text{feed rate of such particles}} \tag{vi}$$

It has been recommended [9] that equation (ii) should not be extrapolated to larger particles as it predicts a non zero value even for those sizes which are not fluidized.

A prior knowledge of the particle size which will be removed from the bed can be obtained by calculating terminal velocities. Stoke's law gives the terminal velocity expression for the laminar flow region as

$$u_t = \frac{(\rho_s - \rho_g)\, g_c D_p^2}{18\,\mu} \quad \text{for} \quad \frac{D_p G}{\mu} < 2.0 \tag{vii}$$

Davidson and Harrison (1971)[1] recommend another correlation for $\left(\frac{D_p G}{\mu}\right) > 2.0$

$$u_t = \frac{0.153\, D_p^{1.14}\, g_c^{0.71}\, (\rho_s - \rho_g)^{0.71}}{\mu^{0.43}\, \rho_g^{0.29}} \qquad \text{(viii)}$$

Particles having terminal velocities less than the superficial gas velocity will be blown out of the system provided the effect of retention by bed is neglected.

2.3 Theoretical Basis of Present Work

From the above discussion, it is observed that estimation of the elutriation rate constant is essential for the design of any elutriation system. Furthermore, the variation of k with the process variables should be confirmed for a continuous system. This work was done to study the relationship between gas velocity, bed height, feed rate and k in a continuous system. The other objective was to compare the elutriation conditions in the continuous mode with that in the batch mode for the same system.

2.4 Other Parameters

The driving force for elutriation is the difference between superficial air velocity and terminal velocity of the particles. Terminal velocity is dependent on particle size and density. Therefore for similar elutriation behaviour of particles of different density, the difference between superficial air velocity and terminal velocity should be equal. The other important factor is attrition. The results obtained for hard particles such as coal char should not be generalised for soft materials.

3. EXPERIMENTAL SETUP AND PROCEDURE

A schematic diagram of the apparatus used for carrying out experiments is shown in figure 5. The three major parts are column, feed and product output system. A sonic sifter was used to obtain the particle size distribution of various samples.

3.1 Fluidization Column

The column used to contain the fluidized bed was of 25 cm inner diameter and 105 cm total height. A large diameter column was chosen to avoid the effect of bed diameter on elutriation rate. Plexiglass was used as the material of construction to have a visual observation of the bed height. A stainless steel perforated plate was placed at a height of 15 cm to support the bed and distribute incoming air. Air was introduced from two side ports below the screen. A product outlet was provided at

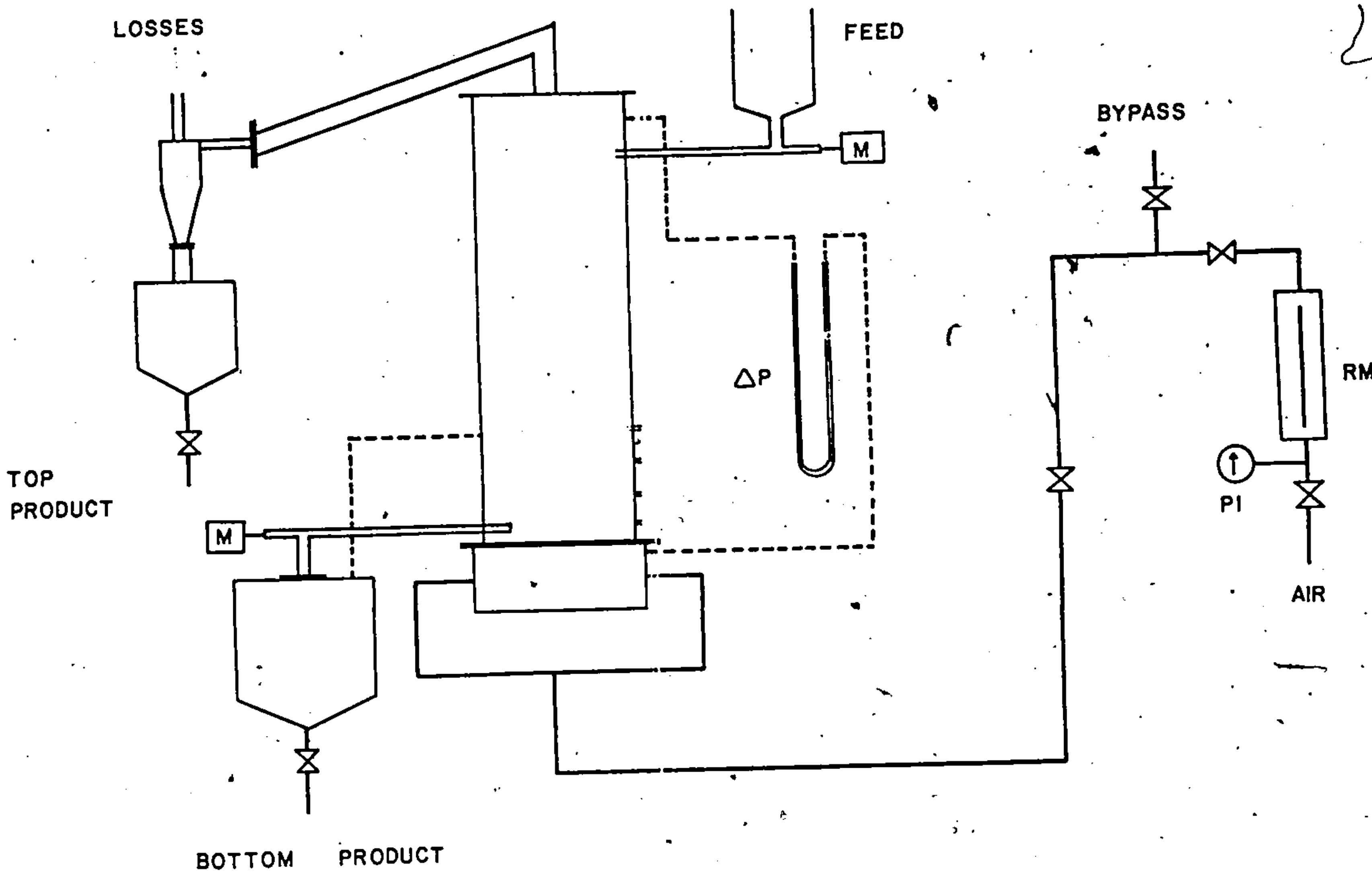


Fig. 5

Schematic diagram of experimental apparatus a height of 4 cm from the screen. Feed was introduced to the column at a height of 70 cm from the screen. The transport disengagement height (TDH) was computed[8] to be 45 cm for a superficial air velocity of 300 cm/min. Thus, the feed point provided sufficient freeboard for the air velocity profile to become uniform. Sample points were provided at a difference of 7.5 cm throughout the height of the column. These could be used to draw out samples if the bed stratification was to be determined. Pressure tap points were provided in the air chamber below the support and near the air outlet. A manometer containing oil of specific gravity 0.827 was connected to these points for measurement of pressure drop across the bed. Entrained material in the air stream was transferred to a cyclone separator through an outlet from the top cover. The pressure drop characteristics for the air distributor is shown in figure 17 of appendix C.

3.2 Cyclone Separator

A cyclone separator was designed to collect the fines as recommended in Perry's handbook[10]. The design basis and exact dimensions are described in appendix C. The material of construction was plexiglass. During the operation it was necessary to tap it intermittently so as to remove the particles sticking on the sides. Bottom products from the cyclone were collected in a chamber and some fines which could not be recovered were thrown out to the atmosphere. These were taken as losses.

3.3 Feed and Output Arrangement

Screw conveyors were used to feed and remove solids. On the feeder side a container filled with feed material was provided. It served two purposes. Firstly, it provided sufficient head to check air from leaking through the screw conveyor and secondly, it avoided frequent manual handling of coal. The screw feeder was rotated with the help of a variable speed "Zero-Max" motor (General Electric, Model No. 5XBHO2FD). Exact dimensions and calibration is given in Appendix C. A similar system was used to remove products from the bed. The material being removed was collected in a plexiglass chamber.

During the initial trial runs, three difficulties were observed which were eliminated by slight modifications. Firstly, solids were settling in the transfer line connecting the column to the cyclone separator. To reduce this amount, it was changed from the horizontal position to an inclined one as shown in the equipment diagram. Also, the total top product was obtained by mixing the amount collected by the cyclone and that settled in the transfer line.

The other operating problem showed up in the bottom product output system. The collection vessel was designed to be at atmospheric pressure while the column had some positive pressure. Due to this pressure difference, some solids were forced out through the screw conveyor even if it was not rotating. This posed a problem in controlling the output rate. It was difficult to check or stop the flow by reducing the speed of rotation. To overcome this difficulty, a pressure balancing line was introduced as shown by the discontinuous line in figure 5. This line connected the collection chamber to the column above the bed surface. To stop the flow of elutriated material through this line, a fine screen was fixed at the opening on the column side. After this modification, the flow was completely controlled by the rotational speed of the screw conveyor.

The third problem came up because of feeding from the top. It was observed that some material was being taken up by air stream directly from the point of entry. To avoid this, the open end of the feeder was moved closer to the column so that the feed point was in the very low velocity region. However, there was a limit to the distance between the feed point and the column wall. It was observed that if this distance was reduced too much, particles started sticking on the wall because of electrostatic forces. Finally it was decided that a distance of 2 cm was reasonable to provide satisfactory feeding.

3.4 Air Flow Line

The air flow line was made of 1 in copper pipe. A pressure gauge manufactured by Binks Manufacturing Co. (Chicago, U.S.A.) was used to measure the pressure of air entering the rotameter. The rotameter, manufactured by Brooks Rotameter Co. (Lansdale, Penn., Type 1110, Serial #39179) was used to measure the input flow rate of air. After the rotameter a bypass line was provided for calibration and adjustment of air flow rate before introducing it to the column. A supply pressure of 5 psig was maintained for air throughout the experiment. The calibration curve for rotameter at 5 psig is given in Appendix C. All the joints in plexiglass portion were flange type with an 'O' ring to stop leaks.

3.5 Experimental Procedure

At the beginning of a run, both the collection vessels, column and transfer line were cleaned of any material in them. The feed material was introduced up to the required bed depth and desired flow rate of air was passed till the bed composition reached a steady state level. The time for steady state attainment was between 2 to 3 hours, depending upon the flow rates. After steady state was reached, again the chambers and transfer line were cleaned. Keeping the air flow and feed rate the same, the run was started. Time for one run was kept close to an hour. When the run was over, bottom and top products were weighed and analyzed. For some experimental runs, the amount of material fed and bed composition at different heights were also measured to observe the actual feed rate and bed stratification. In these runs, a small composition difference was observed at different bed heights showing a weak stratification order. The feed rate actually measured was not much different from the calibrated value. Bed height was maintained at a constant value using output control. For size distribution determination, two samples were frequently analyzed from each product. A total cycle time of 15 min was sufficient for the sonic sifter to provide reproducible results. Screens of sizes 300, 250, 150, 75, 53 and 45 microns were used for determining the size distribution of mixtures.

3.6 Equipment Limitations and Assumptions

(a) Beyond a bed height of 22 cm, it was not possible to conduct any experiments as the pressure in the column was enough to force air through the upper screw feeder.

(b) The sonic sifter gave screen analysis results within a reproducibility error of 5%. The error was mostly due to electrostatic charge on the particles during sifting.

(c) Some losses were experienced in the cyclone. These were estimated on the basis of material balance of particles less than 75 microns.

(d) The error due to electrostatic charge produced on particles during fluidization was taken alongwith other system errors. These errors are shown by the scattering of data points.

(e) A measurement error in the bed height was introduced because of the fluctuation of bed surface. The magnitude of this error could be as high as 5 mm.

(f) Attrition of particles was neglected.

(g) Correction due to shape factor was not taken into account.

4. RESULTS AND DISCUSSION

A summary of all the runs conducted is given in tables 2 through 9 of Appendix A. Three process variables, bed height, air velocity and feed rate were varied independently to study their effect on the elutriation rate constant, k . Run numbers 1 through 14 were conducted for a bed height of 8.5 cm. The air velocity was varied in the range of 115 cm/min to 315 cm/min. Three different feed rates selected were 33.5, 67.0 and 127.5 gm/min. Run numbers 15 through 28 were done for a bed height of 17.5 cm, and 29 through 39 were performed for 22.0 cm bed height under the same operating ranges of air velocity and feed rate. Tables 2 through 4 show the values of the elutriation rate constant, k, obtained in these experiments. A number of batch runs were also done for bed height of 22.0 cm and air velocity ranging between 115 cm/min to 287 cm/min. Table 5 shows the results computed from these batch runs. The results obtained in these runs provide a satisfactory basis for comparison purposes. Tables 6 through 8 show the amount of particles less than

75 microns removed in the top product and residence times of solids and air in the bed. The detailed procedure of calculation is shown in Appendix B for two of the experimental runs.

4.1 Effect of Process Variables on Elutriation Rate Constant

The elutriation rate constant, k , is a strong function of air velocity as shown by equation (ii) in section 2.2. Figures 6 through 9 show this dependence graphically. In these figures, log k has been plotted against log (air velocity) keeping bed height and particle size as parameters. The plots have been prepared from results reported in tables 2 through 5 in Appendix A. Table 1 gives the values of slope and standard deviations of slope for least square fit straight lines. It is obvious from table 1 that variation of k with air velocity is dependent on bed height. As bed height is increased from 8.5 to 22.0 cm, n decreases from 3.3 to 1.86 for particles having average diameter of 64 microns (fraction between 75 and 53 microns). Similar behaviour is observed for other particle sizes. For particles less than 45 microns (average diameter 23 microns) the value of n decreases from 3.61 to 1.63. Hence, it is concluded that as bed height is increased, the effect of air velocity on constant k reduces and the effect is more significant for smaller particles. With respect to the particle size, it is observed

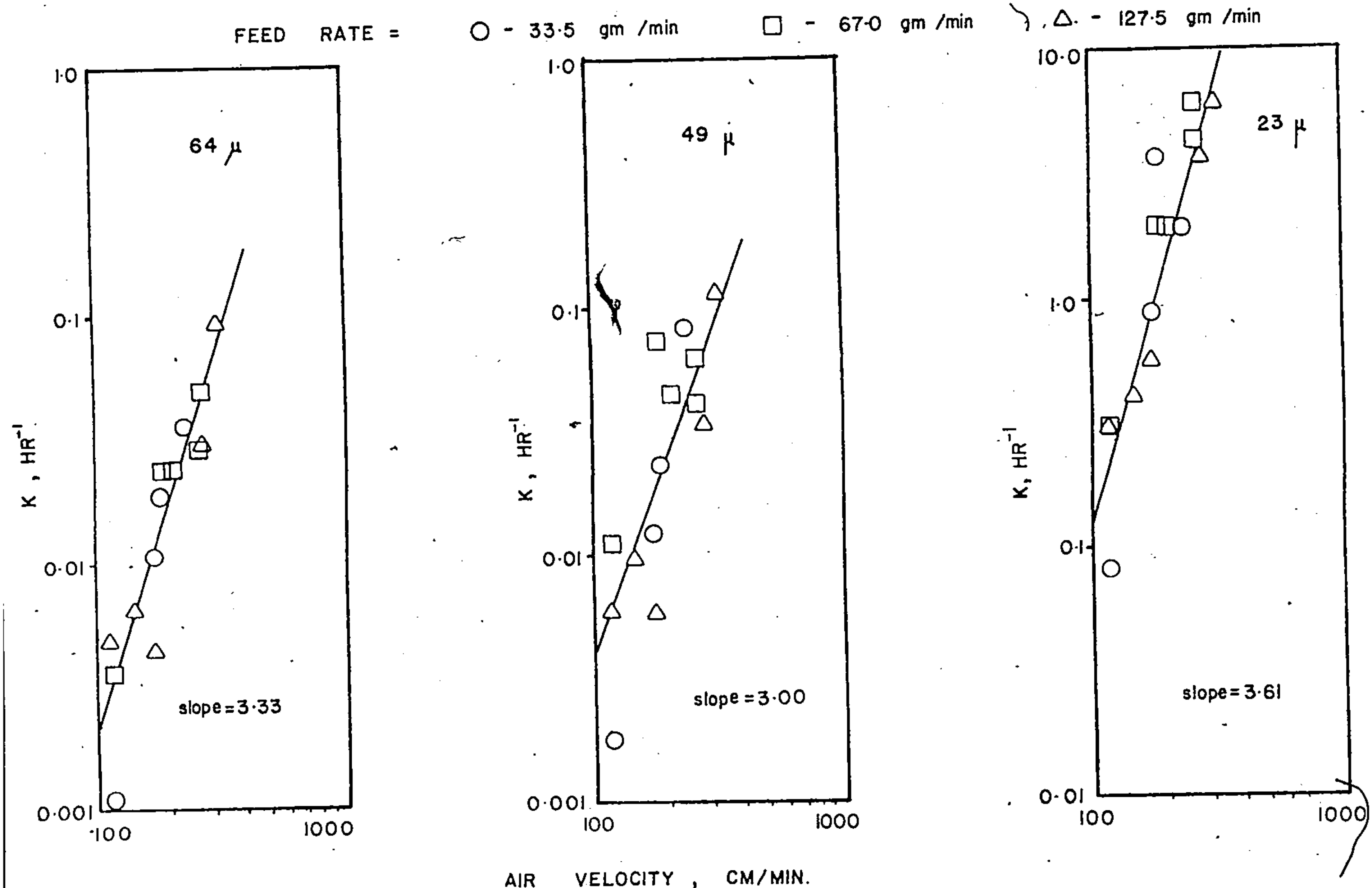


Fig. 6 - Elutriation rate constant for bed height of 8.5 cm

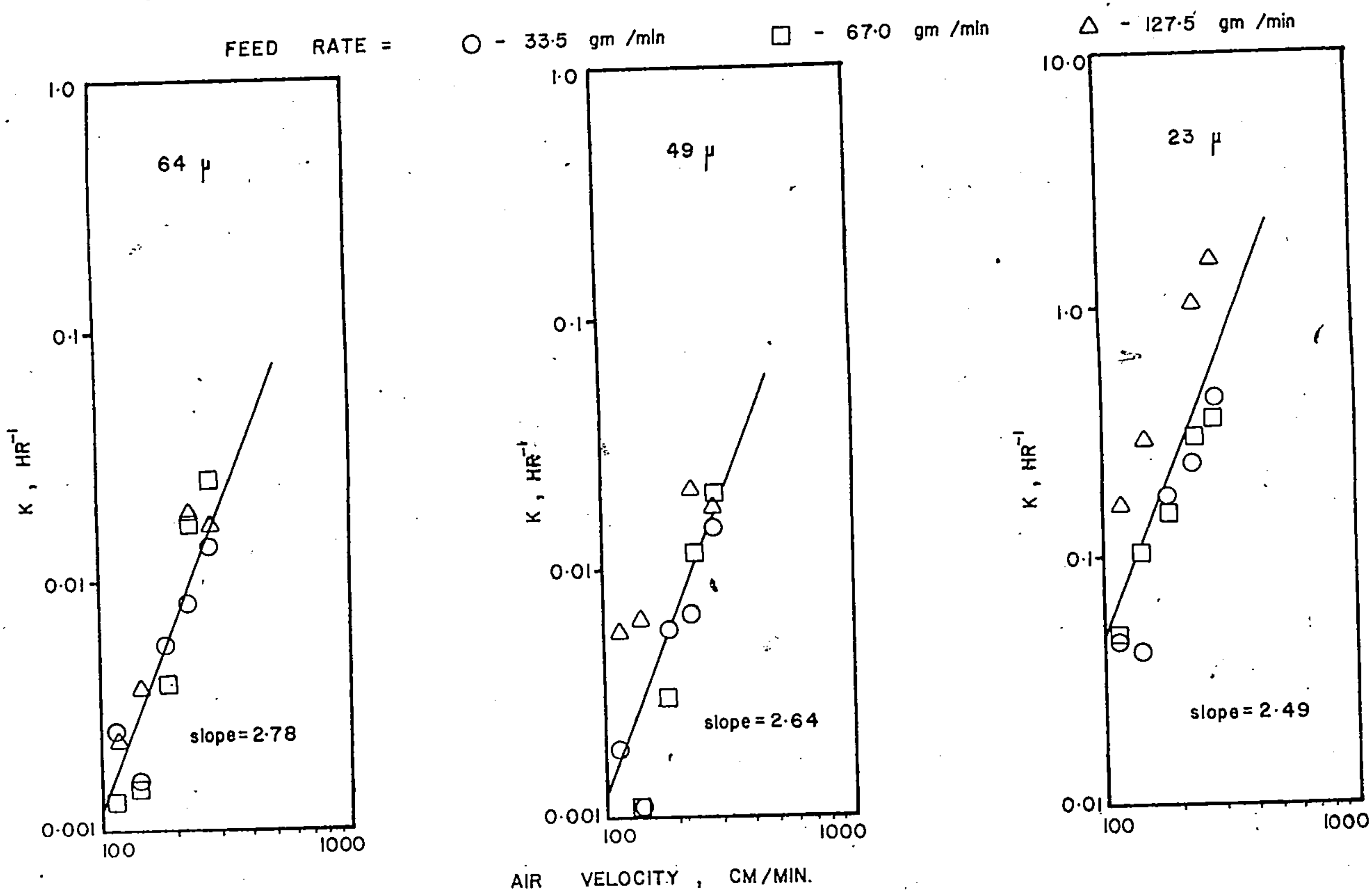


Fig. 7 - Elutriation rate constant for bed height of 17.5 cm

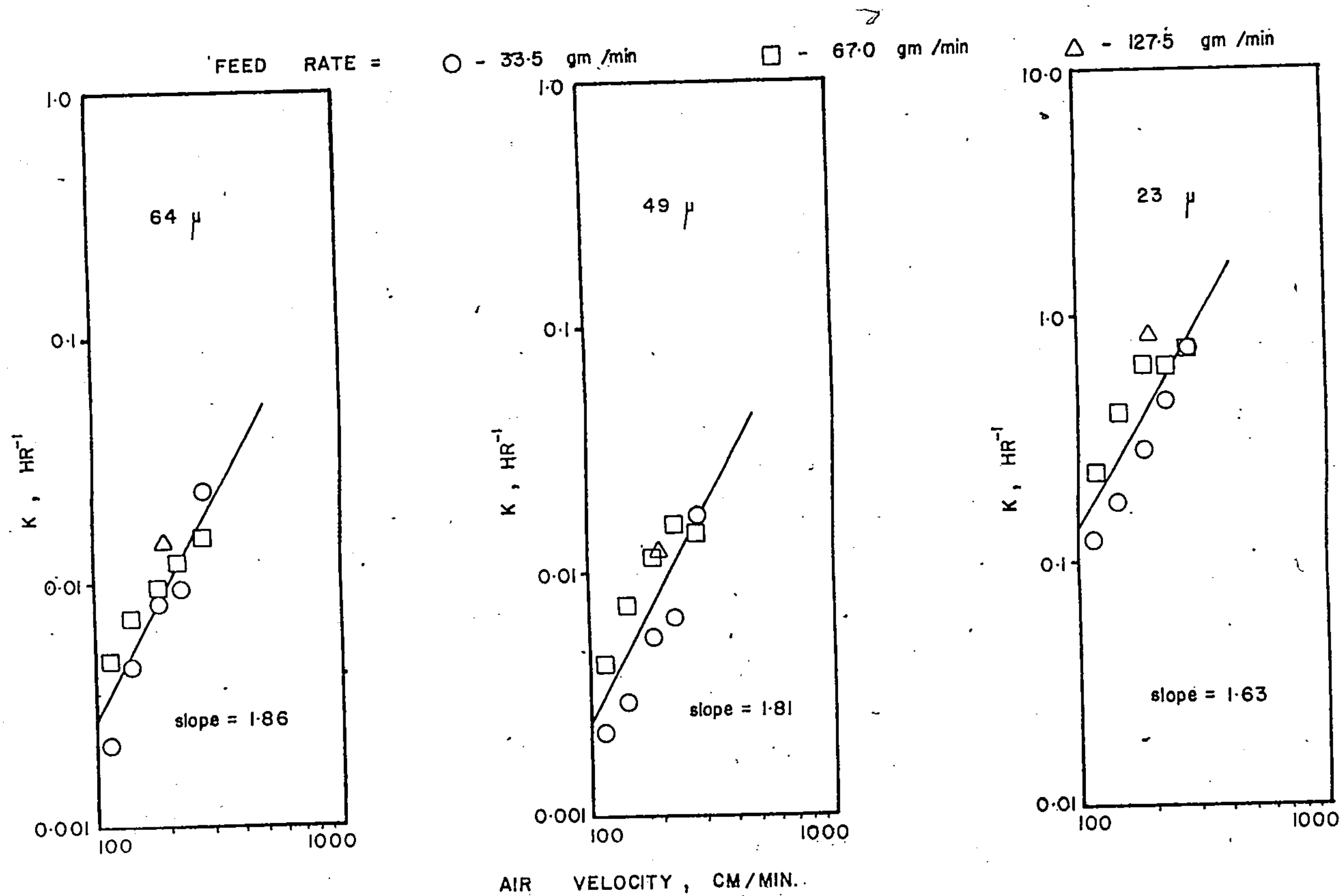
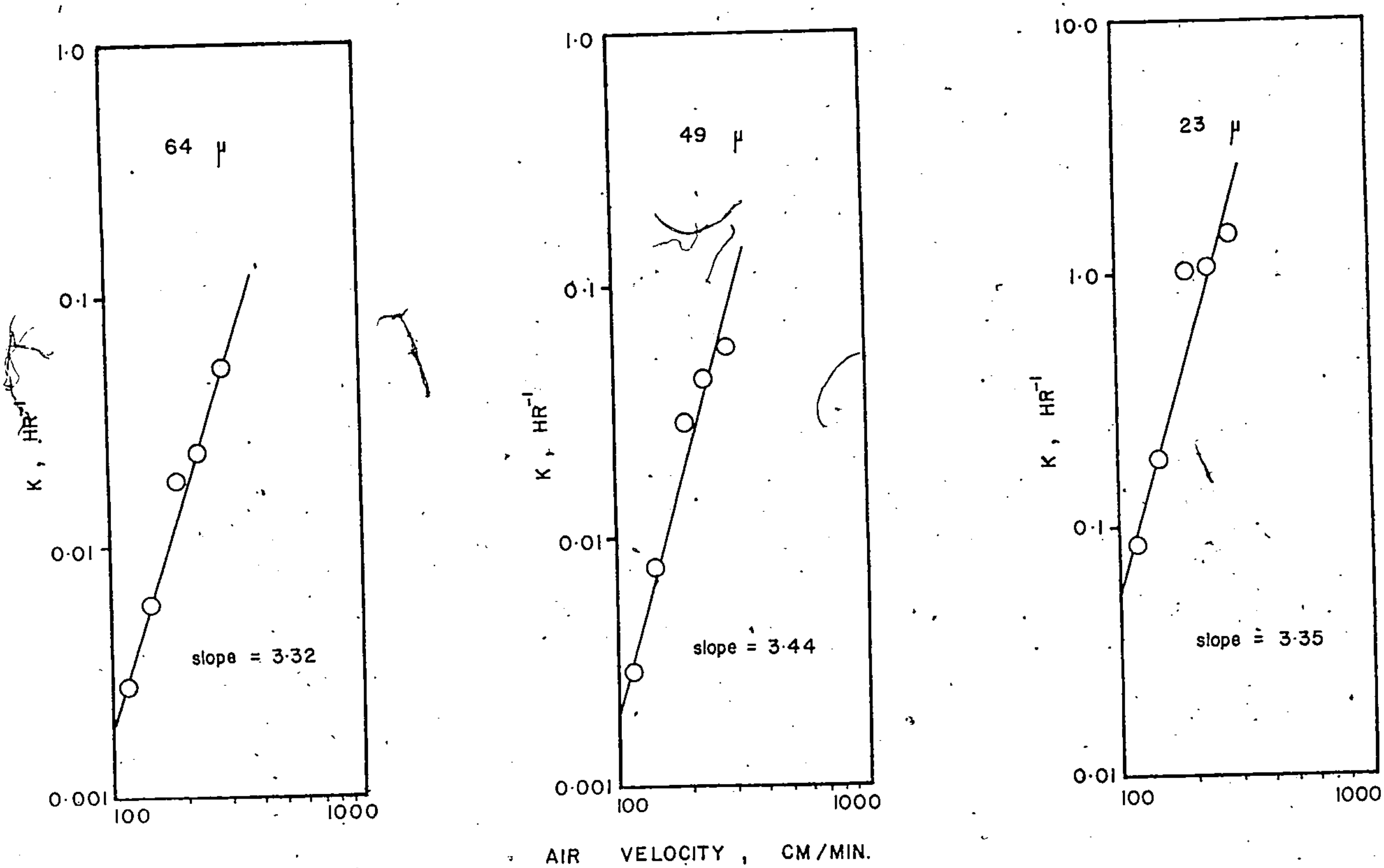


Fig. 8 - Elutriation rate constant for bed height of 22.0 cm



<u>Fig. 9</u> - Elutriation rate constant in batch system for bed height of 22.0 cm

TABLE - 1

Parameters for least square fit equation $\log k = n \log V + B^*$

Bed height (cm)	No. of points	Average particle diameter (microns)								
		64 μ			49 μ			23 μ		
		n	∂_n(%)	$B \times 10^{+9}$	n	∂_n(%)	$B \times 10^{+9}$	n	∂_n(%)	$B \times 10^{+9}$
8.5	14	3.33	13.5	0.38	3.00	20.7	3.24	3.61	13.9	7.94
17.5	14	2.78	12.9	2.69	2.64	21.2	5.50	2.49	22.5	467.73
22.0	11	1.86	16.7	512.86	1.81	23.2	588.84	1.63	25.2	794326.00
22.0 (batch)	5	3.22	7.5	0.66	3.44	14.0	0.28	3.35	20.0	12.30

* k - elutriation rate constant, hr^{-1}
V - air velocity, cm/min
B - intercept

n - slope of the st. line
∂_n - standard deviation in slope

that the effect of air velocity remains of the same order. In the case of batch studies, it is observed that the value of slope or dependence of k on air velocity is nearly the same as for a continuous system in a 8.5 cm bed height. The value of n in batch system is nearly 3.5. In the literature, n has been reported to lie between 2.0 and 4.0 and independent of bed height. In view of the recommendations in the literature[7,9] and data obtained in the present work, it is concluded that for batch systems n is independent of bed height while for steady state systems, it decreases with increasing bed heights.

A theoretical explanation for this difference can be suggested if the mechanism of elutriation is closely followed. It has been mentioned earlier, in section 2.1, that as air bubbles rise in a fluidized bed, an exchange of particles takes place between the wake and dispersed phase. As these bubbles reach the bed surface, the concentration of fines in them is much more than the bed concentration because of stratification of particles in the bed. If a horizontal baffle is introduced in the bed, elutriation rate increases while the effect of a stirrer is to decrease the elutriation rate.

In continuous systems, a similar effect to that of introducing a stirrer is observed. The bed is being fed continuously so it acts as a back mix system. The stratification is reduced because of this continuous feed. Accordingly, air bubbles erupting at the surface are not very much different from the bed in terms of fines concentration. This reduces the elutriation rate of fines as compared to batch runs. Theoretically the change should be much more pronounced for deeper beds, as stratification order is stronger for such beds, as compared to shallow ones in batch systems, and it is expected that the decrease in elutriation rate should be more significant. This expected behaviour is confirmed by the experimental results shown in table 1. For the same height of 22.0 cm, the slope of log k vs. log (air velocity) is approximately half in case of the continuous system as compared to the batch operation.

On the same theoretical grounds, an explanation can be provided for the effect of baffles and stirrers in a batch operation. According to the countercurrent backmixing model (section 2.1) mixing in a bed occurs because of the countercurrent motion of gas and solid particles. When screens are introduced in the bed as horizontal baffles, vertical motion of solids is reduced and upward gas movement becomes more streamlined. Hence, the elutriation rate should increase with baffles. From figure 4, it is observed that the expected behaviour is confirmed experimentally for deeper beds. For smaller bed heights, no effect of baffles is observed. This is expected as the effect of stratification will be more pronounced in deeper beds. Similarly, a stirrer in the bed should decrease the elutriation rate as it reduces stratification order. This is confirmed from the experimental data in figure 3.

4.2 Effect of Bed Height on Elutriation Rate

Figure 10 shows the effect of bed height on the extent of smaller particles removed from the bed. The ratio of particles less than 75 microns diameter present in the overhead product to the amount fed (O/F) is plotted against air velocity keeping bed height and feed rate as parameters. All the points have been plotted taking a basis of one hour run time so that the values on the vertical axis give a direct comparison of the fines removed.

The plots distinctly show that there is a reduction in the amount of fine particles collected as the bed height is increased. Although, as bed height is increased, the freeboard decreases, it is larger than the TDH in all cases and should have no effect. In this case, the factor which governs the removal of smaller particles from the bed is the residence time of air and the extent of fluidization of the bed. The stratification order will of course be larger for deeper bed. However, for continuous systems, this difference is negligible as compared to the difference in air residence time.

4.3 Effect of Residence Time on Elutriation Rate

Figures 11 through 13 show the effect of air residence time on elutriation rate. The graphs have been plotted for feed rates of 33.5, 67.0

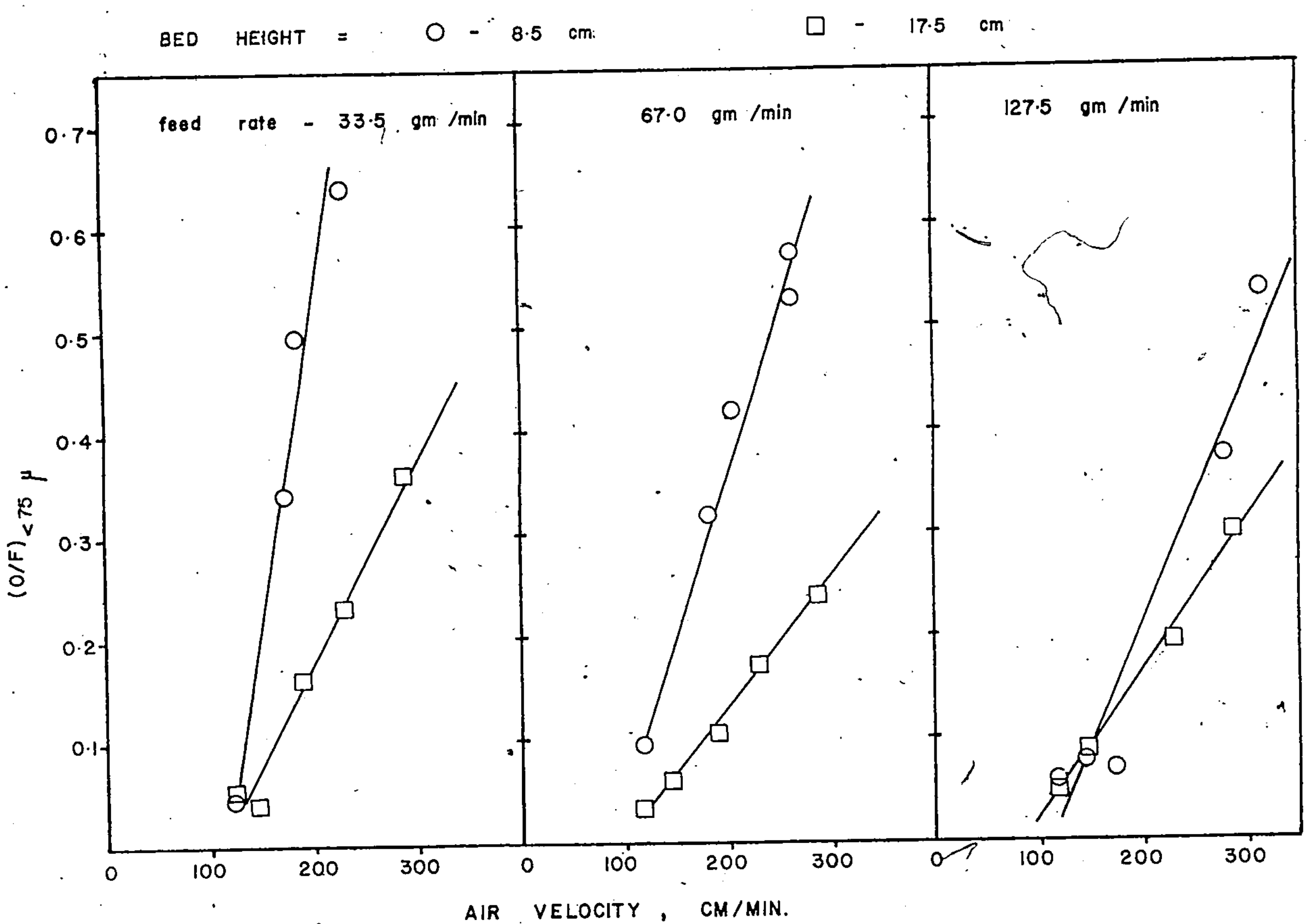


Fig. 10 - Effect of air velocity on fines elutriation

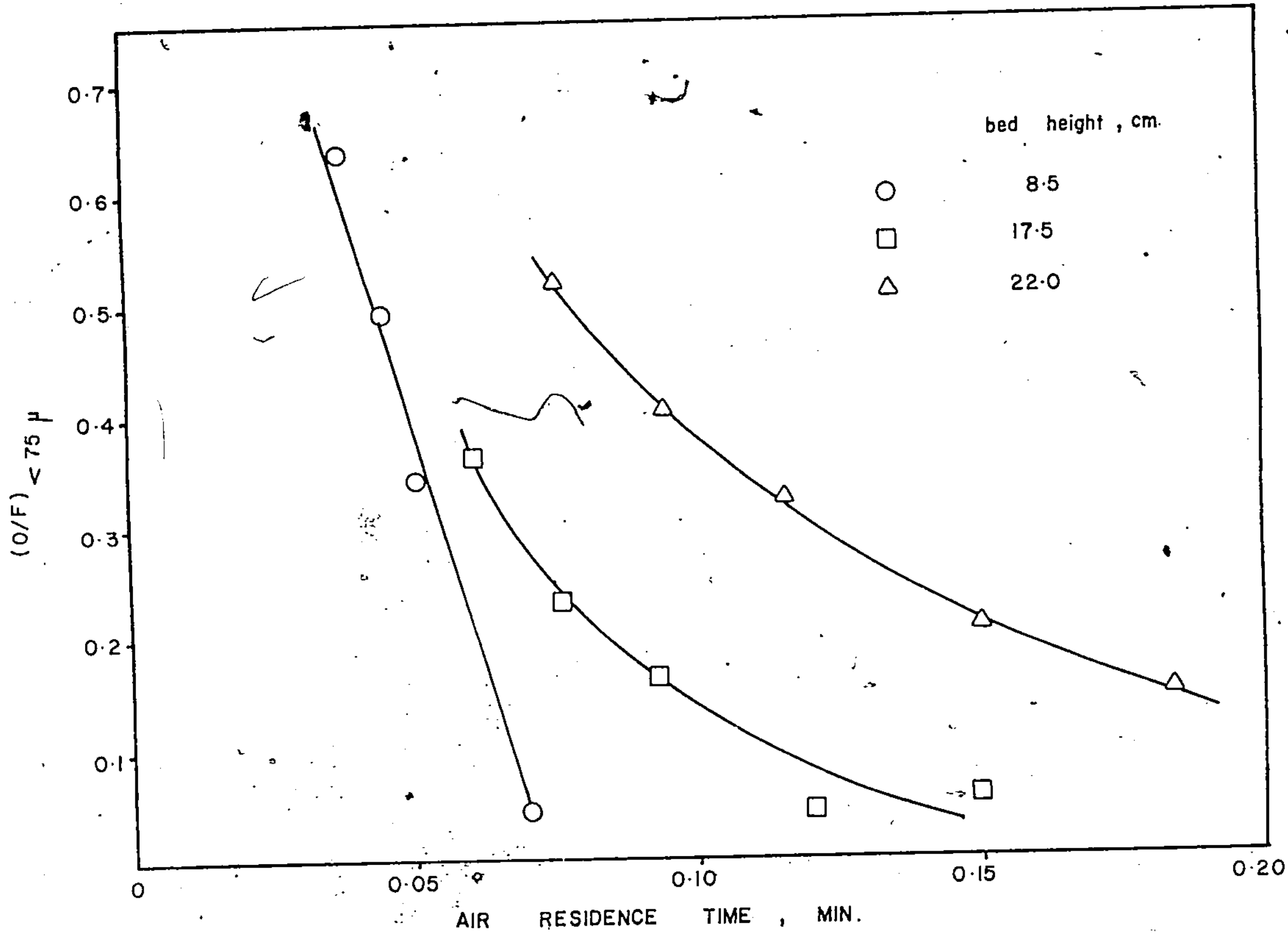


Fig. 11 - Effect of air residence time on fines elutriation (feed rate = 33.5 gm/min)

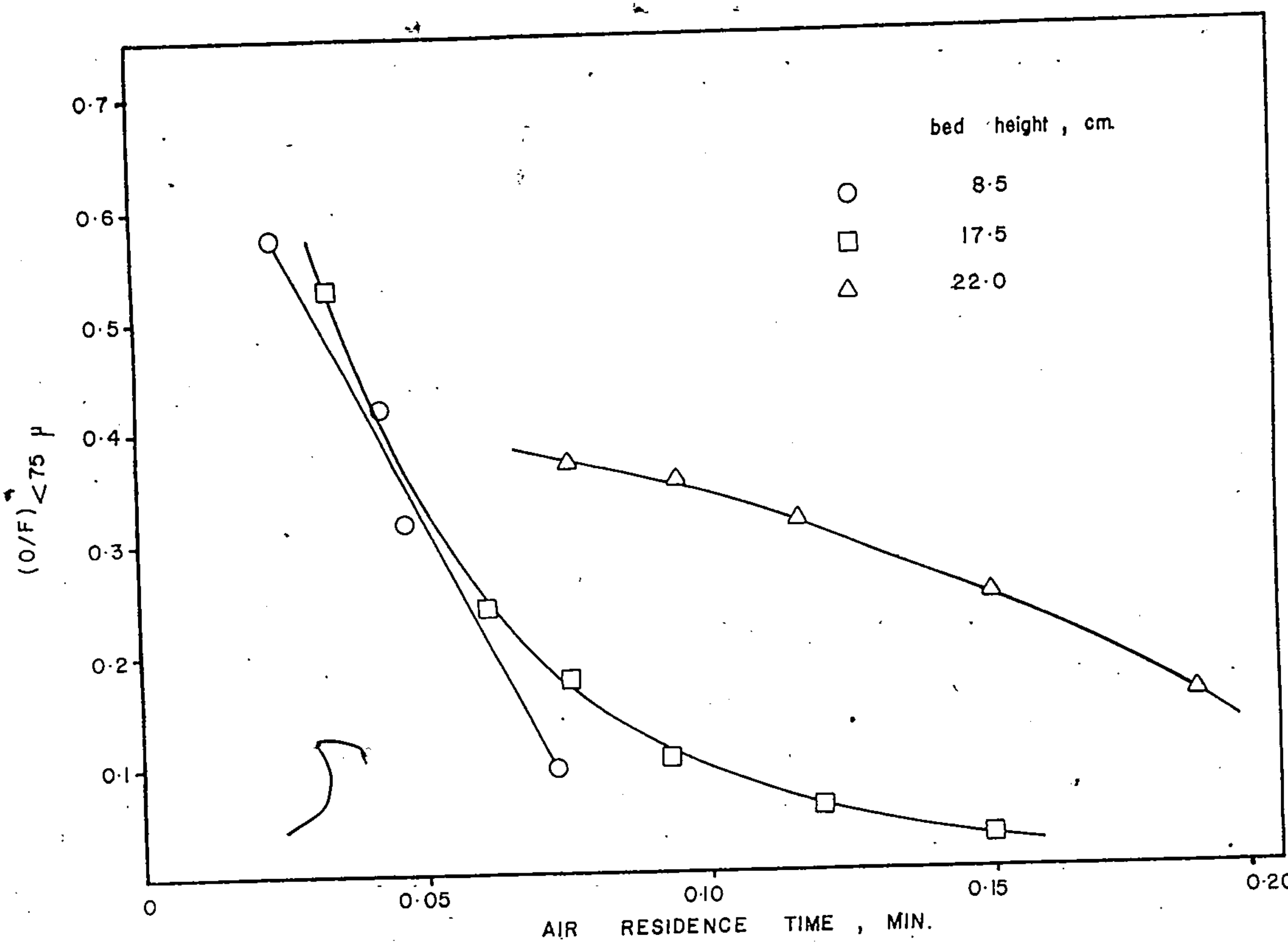


Fig. 12 - Effect of air residence time on fines elutriation
(feed rate = 67.0 gm/min)

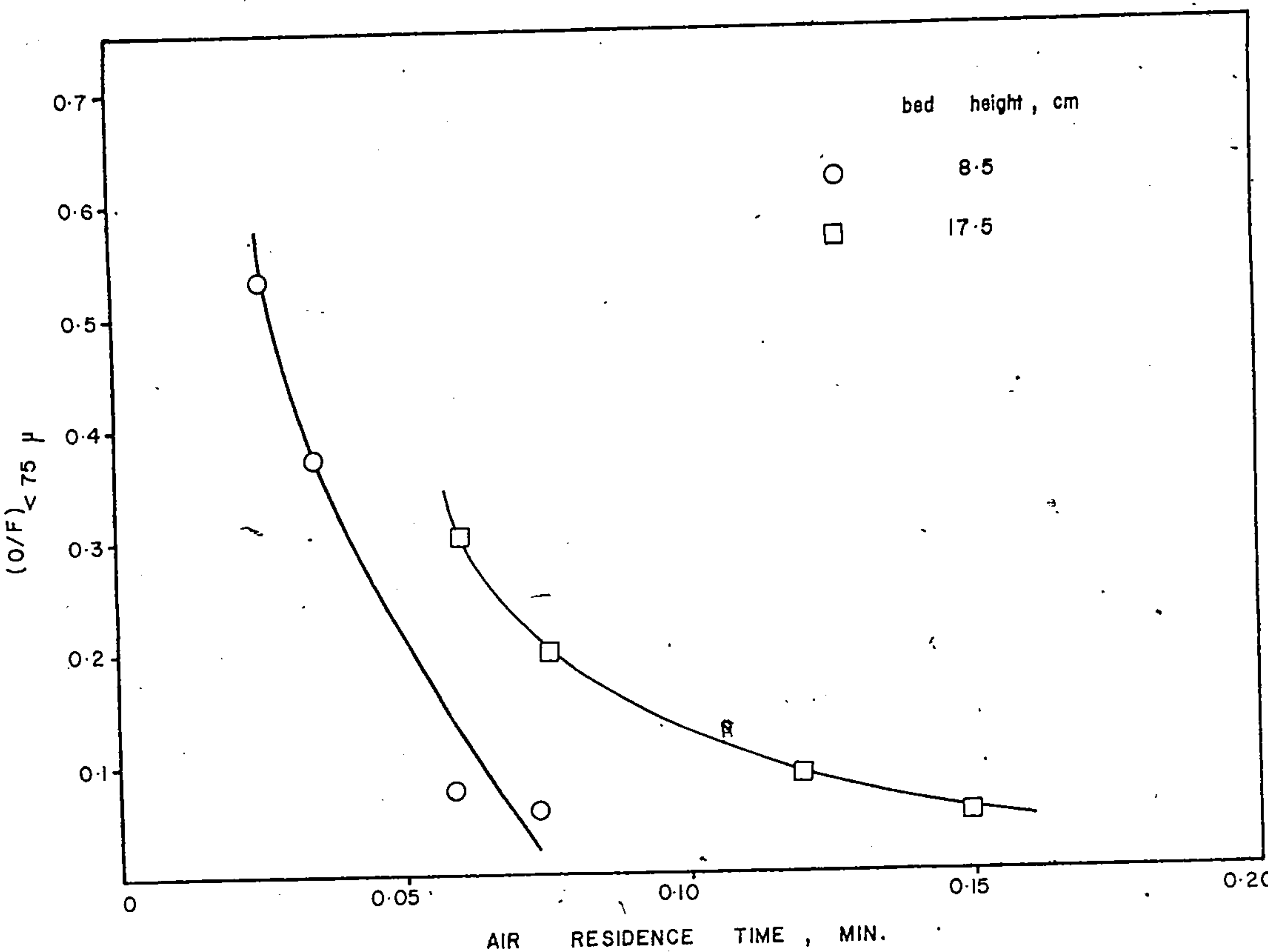


Fig. 13 -- Effect of air residence time on fines elutriation

and 127.5 gm/min respectively. The O/F ratio has been plotted against air residence time keeping bed height as parameter for each feed rate. It is observed that the slope of curves decreases as bed height is increased indicating that the amount of material elutriated is a stronger function of air residence time for lower bed heights.

Figures 14 through 16 have been plotted to determine the effect of solid residence time on elutriation rate constant. Log k has been plotted against log (residence time) for each particle size. It is clear that no conclusions can be drawn out of this plot. The residence time for solids has been calculated using equation (vi) in section 2.2. This equation has been proposed assuming perfect mixing in the bed. However, in the elutriation process, stratification is bound to occur. Due to stratification errors will be introduced in calculating residence time by equation (vi) resulting in the scattering of data shown in figures 14 through 16. Qualitatively, it can be said from these plots that there is a trend showing a decrease in elutriation rate constant with increasing residence time. It is an implicit conclusion as, obviously if k is large for a particle, it will be thrown out of the bed earlier. The large scattering of data points is also due to the fact that residence time is dependent on several factors such as air velocity, bed height, feed rate, particle size, electrostatic forces, etc. To study the effect properly, factorially designed experiments should be performed, taking sufficient number of experimental

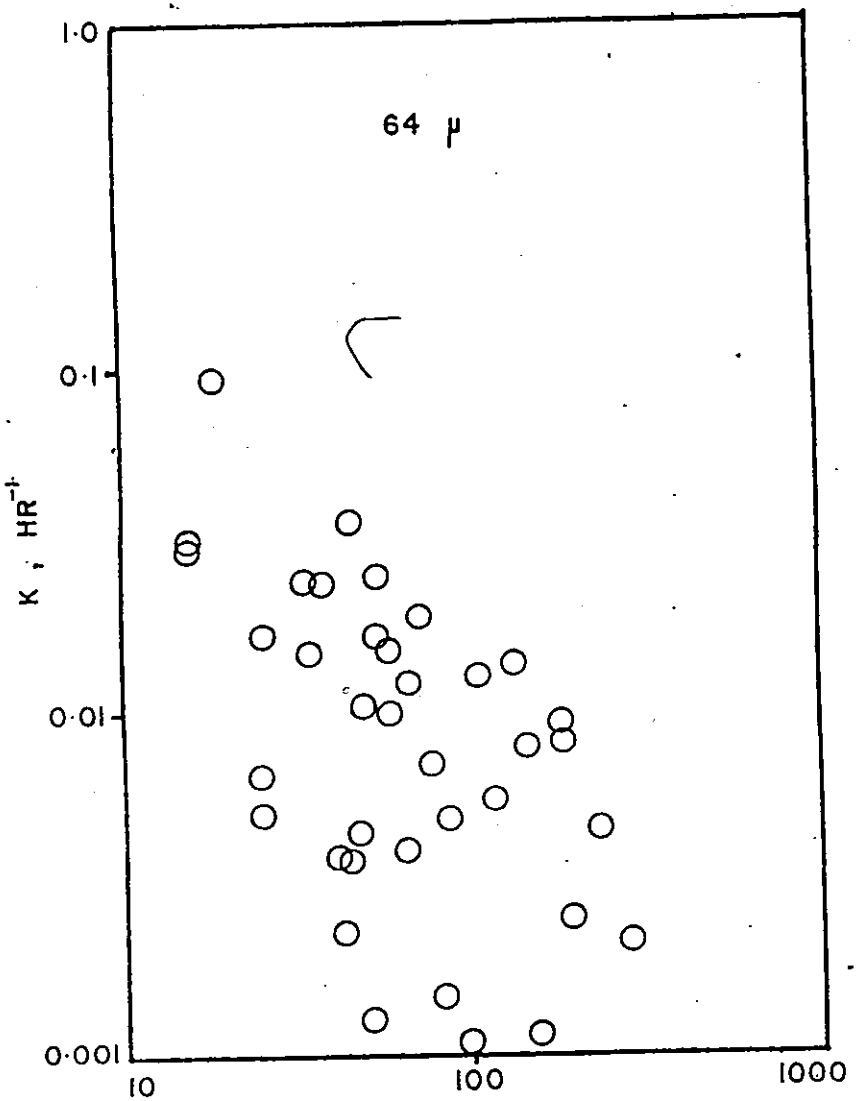


Fig. 14

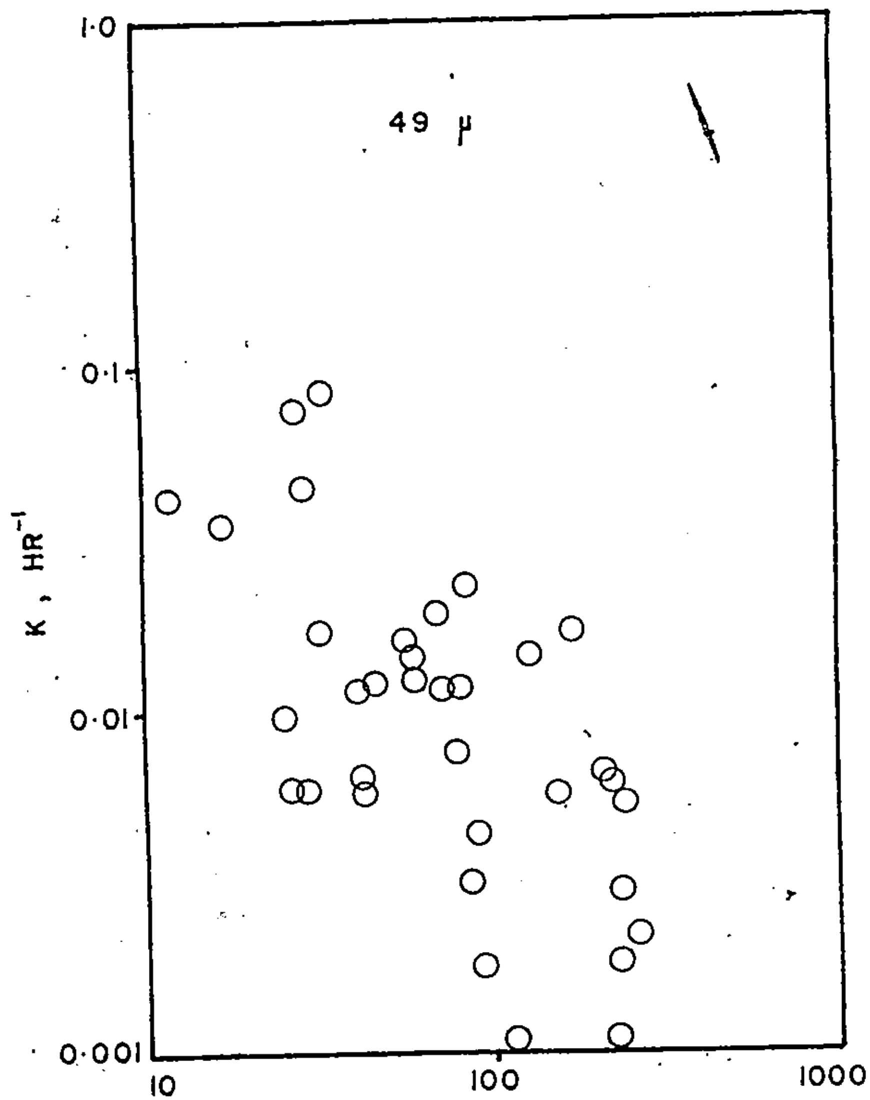


Fig. 15

Effect of solid residence time on elutriation rate constant

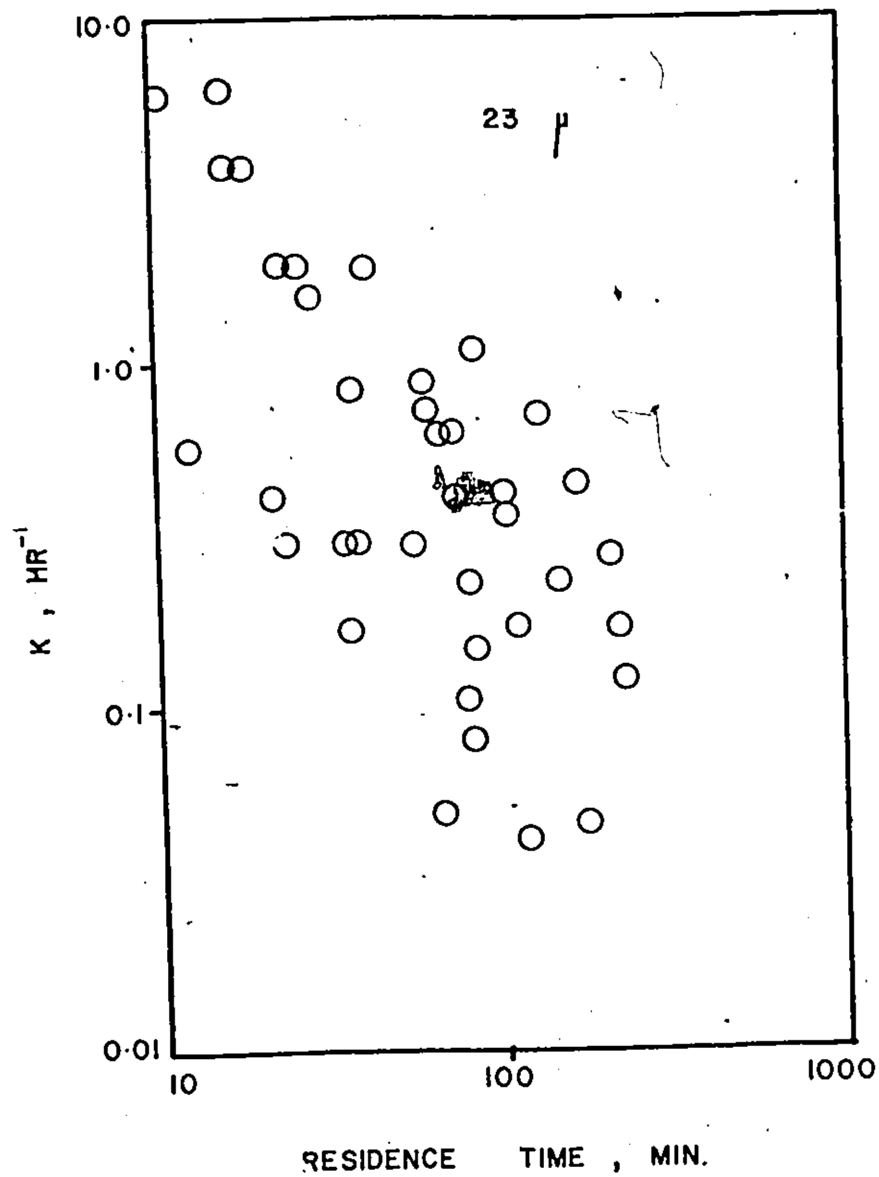


Fig. 16

Effect of solid residence time on elutriation rate constant points. So, the final conclusion is that no satisfactory results can be drawn for the effect on k due to solids residence time using the data available in this work, only very general trends can be observed.

4.4 Stratification of Particles in Bed

In several runs on continuous elutriation, samples were drawn from the bed at two different heights and were analysed to observe variation of fines concentration along the bed height. Table 9 in Appendix A shows the results obtained for these runs. Studies were conducted for 22.0 cm bed height runs where maximum stratification was expected.

It was observed that very small build up of fines occurred at the surface as compared to batch data reported in section 2.1. As the air velocity is increased, the stratification order reduces. Thus, it confirms that treatment of the continuous elutriation scheme as a back mix system should be fairly consistent.

5. CONCLUSION

1. The effect of air velocity on elutriation rate constant decreases as the bed height is increased for a continuous system. The dependence of constant k on air velocity is

$$k \propto (\text{air velocity})^n$$

and the factor n for 23 micron particle size varies from 3.61 to 1.63 as the bed height is increased from 8.5 cm to 22.0 cm. Similar behaviour was observed for other particle sizes.

2. The elutriation rate constant dependence on air velocity decreases as bed height is increased in a continuous system in contrast to batch systems. Therefore, for higher bed heights, a batch process is better in performance. However, around the bed height of 8.5 cm (bed height to bed diameter ratio nearly 1/3) the elutriation rate constant for a batch or a continuous system becomes equal.

3. In addition to bed height, particle diameter and air velocity, the effect of stratification of the fluidized bed is quite significant. For the design of any elutriation system this effect must be taken into account.

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APPENDIX A

SUMMARY OF EXPERIMENTAL DATA

Feed Material

The composition of feed material for most of the runs was as given below :

Size (microns)	wt. %
+300	0.571
-300 + 250	0.427
-250 + 150	21.921
-150 + 75	35.221
- 75 + 53	12.230
- 53 + 45	9.771
- 45	19.913

Physical Properties

Density of coal particles	= 1.5 gm/cc
Bed density	= 0.823 gm/cc
Air viscosity	= 0.018 cp
Bed diameter	= 24.94 cm

Terminal velocity for (75-53) microns	= 166.4 cm/min
(53-45) microns	= 115.6 cm/min
less than 45 microns	= 91.1 cm/min

TABLE - 2

k constant for bed height ~ 8.5 cm

Run No.	Feed Rate (gm /min)	Air Velocity (cm /min)	k constant (hr^{-1}) 64 μ	49 μ	22 μ
1.	33.5	117.7	0.00110	0.00177	0.08144
2.	33.5	174.0	0.01083	0.01251	0.87742
3.	33.5	184.8	0.01906	0.02373	3.72495
4.	33.5	230.3	0.03654	0.08433	1.90028
5.	67.0	117.7	0.00372	0.01113	0.30522
6.	67.0	182.0	0.02460	0.07494	1.97161
7.	67.0	204.7	0.02458	0.04496	1.97399
8.	67.0	261.6	0.05026	0.06526	4.21095
9.	67.0	264.4	0.03027	0.04118	6.13993
10.	127.5	117.1	0.00490	0.00591	0.30723
11.	127.5	145.6	0.00658	0.00954	0.41641
12.	127.5	174.0	0.00446	0.00574	0.57215
13.	127.5	278.6	0.03136	0.03380	3.71271
14.	127.5	315.6	0.09498	0.11204	6.31400

TABLE - 3

k constant for bed height ~ 17.5 cm

Run No.	Feed Rate (gm /min)	Air Velocity (cm /min)	k constant (hr^{-1})		
			64 μ	49 μ	22 μ
15.	33.5	117.1	0.00250	0.00188	0.04635
16.	33.5	145.6	0.00161	0.00113	0.04186
17.	33.5	188.2	0.00554	0.00568	0.17429
18.	33.5	230.3	0.00798	0.00654	0.23617
19.	33.5	287.1	0.01338	0.01430	0.42962
20.	67.0	117.1	0.00134	0.00087	0.04873
21.	67.0	145.6	0.00149	0.00108	0.10583
22.	67.0	188.2	0.00390	0.00311	0.15031
23.	67.0	230.3	0.01724	0.01155	0.29951
24.	67.0	287.1	0.02518	0.01944	0.35809
25.	127.5	117.1	0.00234	0.00567	0.16648
26.	127.5	145.6	0.00378	0.00642	0.30281
27.	127.5	230.3	0.01853	0.02178	1.00997
28.	127.5	287.1	0.01710	0.01741	1.57146

TABLE - 4

k constant for bed height ~ 22.0 cm

Run No.	Feed Rate (gm /min)	Air Velocity (cm /min)	k constant (hr^{-1})		
			64 µ	49 µ	22 µ
29.	33.5	117.1	0.00224	0.00221	0.12221
30.	33.5	145.6	0.00461	0.00292	0.17107
31.	33.5	188.8	0.00820	0.00536	0.28319
32.	33.5	230.3	0.00940	0.00636	0.45016
33.	33.5	287.1	0.02390	0.01707	0.71590
34.	67.0	117.1	0.00488	0.00427	0.23425
35.	67.0	145.6	0.00716	0.00752	0.40709
36.	67.0	188.2	0.00966	0.0117[illegible]	0.64050
37.	67.0	230.3	0.01226	0.01589	0.62748
38.	67.0	287.1	0.01563	0.01456	0.73509
39.	127.5	191.6	0.01500	0.01217	0.83647

TABLE - 5

k constants for batch study ~ 22.0 cm bed

Run No.	Air Velocity (cm /min)	Initial Bed h t. cm		Final Bed h t. cm		k constant (hr $^{-1}$)		
		Settled	Fluidized	Settled	Fluidized	64 μ	49 μ	22 μ
40.	117.1	15.2	21.8	15.0	21.3	0.00280	0.00289	0.08413
41.	145.6	14.7	21.8	14.2	21.1	0.00601	0.00756	0.18524
42.	188.8	14.2	21.8	12.2	18.8	0.01828	0.02829	1.03253
43.	230.3	13.7	21.8	11.4	18.3	0.02370	0.04301	1.07607
44.	287.1	13.2	21.8	10.4	17.3	0.05201	0.05678	1.44920

TABLE - 6

Residence times for bed height ~ 8.5 cm

Run No.	Feed Rate (gm /min)	Air Velocity cm /min	Gas Residence time (min)	Overhead/Feed for < 64 μ	Solid Residence time(min)		
					64 μ	49 μ	22 μ
1.	33.5	117.7	0.071	0.045	97.7	94.0	83.7
2.	33.5	174.0	0.051	0.343	50.3	61.3	60.2
3.	33.5	184.8	0.045	0.496	72.9	87.4	16.5
4.	33.5	230.3	0.038	0.637	46.2	33.4	41.0
5.	67.0	117.7	0.073	0.095	45.3	42.2	38.1
6.	67.0	182.0	0.047	0.317	38.2	27.9	24.1
7.	67.0	204.7	0.043	0.420	34.1	29.0	26.3
8.	67.0	261.6	0.034	0.527	16.3	15.6	9.5
9.	67.0	264.4	0.034	0.573	11.6	11.9	10.9
10.	127.5	117.1	0.074	0.059	24.8	26.6	23.6
11.	127.5	145.6	0.059	0.075	24.5	25.5	22.3
12.	127.5	174.0	0.051	0.069	27.0	29.5	12.7
13.	127.5	278.6	0.035	0.372	15.8	17.0	18.6
14.	127.5	315.6	0.026	0.532	8.9	7.5	16.6

TABLE - 7

Residence times for bed height ~ 17.5 cm

Run No.	Feed Rate (gm /min)	Air Velocity (cm /min)	Air Residence time (min)	Overhead/Feed for < 64 μ	Solid Residence time(min)		
					64 μ	49 μ	22 μ
15.	33.5	117.1	0.150	0.055	198.3	234.0	174.6
16.	33.5	145.6	0.120	0.042	158.3	228.5	119.1
17.	33.5	188.2	0.093	0.163	118.3	153.5	113.5
18.	33.5	230.3	0.076	0.231	146.3	211.5	148.2
19.	33.5	287.1	0.061	0.361	108.3	134.6	102.0
20.	67.0	117.1	0.150	0.032	50.9	82.4	66.7
21.	67.0	145.6	0.120	0.059	83.3	118.4	80.8
22.	67.0	188.2	0.093	0.103	65.8	87.6	84.4
23.	67.0	230.3	0.076	0.174	55.0	73.0	57.5
24.	67.0	287.1	0.061	0.239	54.5	71.7	105.0
25.	127.5	117.1	0.150	0.050	43.5	43.0	36.9
26.	127.5	145.6	0.120	0.088	41.4	43.4	35.9
27.	127.5	230.3	0.076	0.194	46.5	42.0	23.2
28.	127.5	287.1	0.061	0.300	25.8	32.2	28.9

TABLE - 8

Residence times for bed height ~ 22.0 cm

Run No.	Feed Rate (gm /min)	Air Velocity (cm /min)	Air Residence time (min)	Overhead/Feed for < 64 μ	Solid Residence time(min)		
					64 μ	49 μ	22 μ
29.	33.5	117.1	0.184	0.149	288.3	262.3	226.0
30.	33.5	145.6	0.150	0.207	232.4	238.1	223.5
31.	33.5	188.8	0.116	0.322	179.9	244.9	210.3
32.	33.5	230.3	0.095	0.404	182.7	218.2	166.1
33.	33.5	287.1	0.076	0.519	138.7	171.5	132.4
34.	67.0	117.1	0.186	0.156	84.3	91.1	82.1
35.	67.0	145.6	0.150	0.245	78.0	79.3	74.6
36.	67.0	188.2	0.116	0.354	88.9	82.3	76.0
37.	67.0	230.3	0.095	0.370	66.3	58.0	69.8
38.	67.0	287.1	0.076	0.315	58.1	59.5	62.2
39.	127.5	191.6	0.114	0.257	34.9	47.3	38.0

TABLE - 9

Stratification data for bed height ~ 22.0 cm

Particle size μ	Run No. 34		Run No. 37		Run No. 38		Run No. 29	
	12.0 cm	22.0 cm	12.0 cm	22.0 cm	12.0 cm	22.0 cm	12.0 cm	22.0 cm
+300	0.536	0.568	0.469	0.568	0.527	0.485	0.621	0.752
-300+250	0.629	0.673	0.380	0.477	0.632	0.523	0.576	0.428
-250+150	20.382	18.415	21.953	18.898	20.628	19.723	19.103	19.507
-150+ 75	36.437	32.071	35.020	35.644	37.431	35.341	35.956	31.323
- 75+ 53	12.069	10.349	12.301	10.961	10.726	12.459	10.959	11.576
- 53+ 45	10.284	11.436	8.559	7.581	10.054	9.682	11.532	9.938
- 45	19.663	26.558	21.317	25.870	20.002	21.787	21.253	26.476
	Run No. 30		Run No. 31		Run No. 32		Run No. 33	
	12.0 cm	22.0 cm	12.0 cm	22.0 cm	12.0 cm	22.0 cm	12.0 cm	22.0 cm
+300	0.536	0.681	0.635	0.781	0.572	0.653	0.603	0.734
-300+250	0.387	0.529	0.576	0.392	0.428	0.723	0.614	0.626
-250+150	22.221	17.993	21.668	19.419	73.981	22.862	22.595	22.886
-150+ 75	34.119	31.416	33.219	34.122	33.113	32.149	34.105	35.716
- 75+ 53	12.146	11.943	13.653	12.457	10.179	12.432	12.320	11.618
- 53+ 45	10.524	11.056	8.956	10.308	11.513	8.653	11.241	8.293
- 45	21.067	26.382	21.793	22.521	20.214	22.528	19.522	20.127

For flow rates see Table 8.

APPENDIX B

CALCULATIONS FOR RUN NOS. 25 and 40

Run No. 25

Experimental observations :

Feed rate	= 127.5 gm/min	(scale reading = 10.0)
Air flow rate	= 2.02 cu ft/min	(scale reading = 2.9)
Fluidized bed ht.	= 6.9 in	= 17.53 cm
Settled bed ht.	= 5.1 in	= 12.95 cm
ΔP	= 3.34 in of oil	(S.G = 0.827)
Run time	= 60 min	
Wt. of overhead product	= 58 gm	
Wt. of bottom product	= 7364 gm	
Wt. of feed	= 7620 gm	

Screen Analysis :

No.	Size	wt. % Feed	Top	Bottom
1.	+300	0.521	0.648	0.529
2.	-300 + 250	0.422	0.653	0.714
3.	-250 + 150	21.921	0.660	23.068
4.	-150 + 75	35.221	1.643	34.543
5.	- 75 + 53	12.230	0.952	12.978
6.	- 53 + 45	9.771	1.821	10.241
7.	- 45	19.913	93.623	17.927

Calculations

Area of bed $= \pi \times \frac{(24.94)^2}{4} = 488.27 \ cm^2$

Air flow rate $= 2.02 \times (30.48)^3 = 57200$ cc/min

Superficial air velocity $= 57200 / 488.27 = 117.13$ cm/min

Feed rate $= 7620 / 60.0 = 127.0$ gm/min

Wt. of bed material $= 488.27 \times 0.823 \times 12.95 = 5204$ gm

Wt. % of fines ($< 75\mu$) in feed $= 12.23 + 9.771 + 19.913 = 41.914$

Wt. of fines fed $= 7620 \times 0.41914 = 3194$ gm

Wt. % of fines in overhead $= 0.952 + 1.821 + 93.623 = 96.396$

Wt. of fines in overhead product $= 58 \times 0.96396 = 56$ gm

Wt. % of fines in bottom $= 12.978 + 10.241 + 17.927 = 41.146$

Wt. of fines in bottom product $= 7364 \times 0.41146 = 3030$ gm

Losses in cyclone $= 3194 - 56 - 3030 = 108$ gm

Total overhead product $= 108 + 58 = 166$ gm

$$\frac{\text{Amt. of fines in overhead}}{\text{Amt. of fines in feed}} = \frac{166 \times 0.96396}{7620 \times 0.41914} = \frac{160}{3194} = 0.050$$

Calculation of k constant -

$$k_i = \frac{O \times y_i}{V \times x_i} \qquad \text{from equation (v)}$$

i - particle size

O - overhead product, gm/hr

V - bed amount, gm

y - top concentration

x - bed concentration

for particle size (-75 + 53) μ

$$k = \frac{166 \times 0.952}{5204 \times 12.978} = 0.00234 \ \mathrm{hr}^{-1}$$

for particle size (-53 + 45) μ

$$k = \frac{166 \times 1.821}{5204 \times 10.241} = 0.00567 \ \mathrm{hr}^{-1}$$

for particle size (-45) μ

$$k = \frac{166 \times 93.623}{5204 \times 17.927} = 0.16648 \ \mathrm{hr}^{-1}$$

Residence time calculation -

$$\text{Residence time of air} = \frac{17.53}{117.13} = 0.150 \text{ min}$$

Residence time of solids

$$\bar{t}_i = \frac{V \times X_i}{F \times Z_i} \qquad \text{from equation (vi)}$$

i - particle size

V - bed amount, gm

F - feed rate, gm/min

X - bed concentration

Z - feed concentration

for particle size (-75 + 53) μ

$$\bar{t} = \frac{5204 \times 12.978}{127.0 \times 12.230} = 43.5 \text{ min}$$

for particle size (-53 + 45) μ

$$\bar{t} = \frac{5204 \times 10.241}{127.0 \times 9.771} = 43.0 \text{ min}$$

for particle size (-45) μ

$$\bar{t} = \frac{5204 \times 17.927}{127.0 \times 19.913} = 36.9 \text{ min}$$

Run No. 40

Experimental observations :

Initial settled bed ht.	= 6.0 in = 15.2 cm
Initial fluidized bed ht.	= 8.6 in = 21.8 cm
Final settled bed ht.	= 5.9 in = 15.0 cm
Final fluidized bed ht.	= 8.4 in = 21.3 cm
Air flow rate	= 2.06 cu ft/min (scale reading = 3.0)
Run time	= 60 min
Pressure drop	= 4.0 in average
Top product	= 52 gm

Screen Analysis :

No.	Size	Initial (wt. %)	Final (wt. %)	Top (wt. %)
1.	+300	0.521	0.295	0.334
2.	-300 + 250	0.422	0.287	0.346
3.	-250 + 150	21.921	21.269	1.623
4.	-150 + 75	35.221	35.876	2.068
5.	- 75 + 53	12.230	11.844	2.006
6.	- 53 + 45	9.771	13.768	2.022
7.	-45	19.913	16.659	91.600

Calculations

Initial wt. of material in bed = $488.27 \times 0.823 \times 15.2 = 6124$ gm

Final wt. of material in bed = $488.27 \times 0.823 \times 15.0 = 6022$ gm

Wt. of overhead product = 102 gm

Average bed concentration -

$(-75 + 53)\ \mu$ $= \dfrac{11.844 + 12.230}{2} = 12.037$

$(-53 + 45)\ \mu$ $= \dfrac{13.768 + 9.771}{2} = 11.770$

$(-45)\ \mu$ $= \dfrac{16.659 + 19.913}{2} = 18.286$

Average bed weight $= \dfrac{6124 + 6022}{2} = 6073$ gm

k constants -

$(-75 + 53)\ \mu$ $= \dfrac{102 \times 2.006}{6073 \times 12.037} = 0.00280\ hr^{-1}$

$(-53 + 45)\ \mu$ $= \dfrac{102 \times 2.022}{6073 \times 11.770} = 0.00289\ hr^{-1}$

$(-45)\ \mu$ $= \dfrac{102 \times 91.600}{6073 \times 18.286} = 0.08413\ hr^{-1}$

APPENDIX C

EQUIPMENT DETAILS

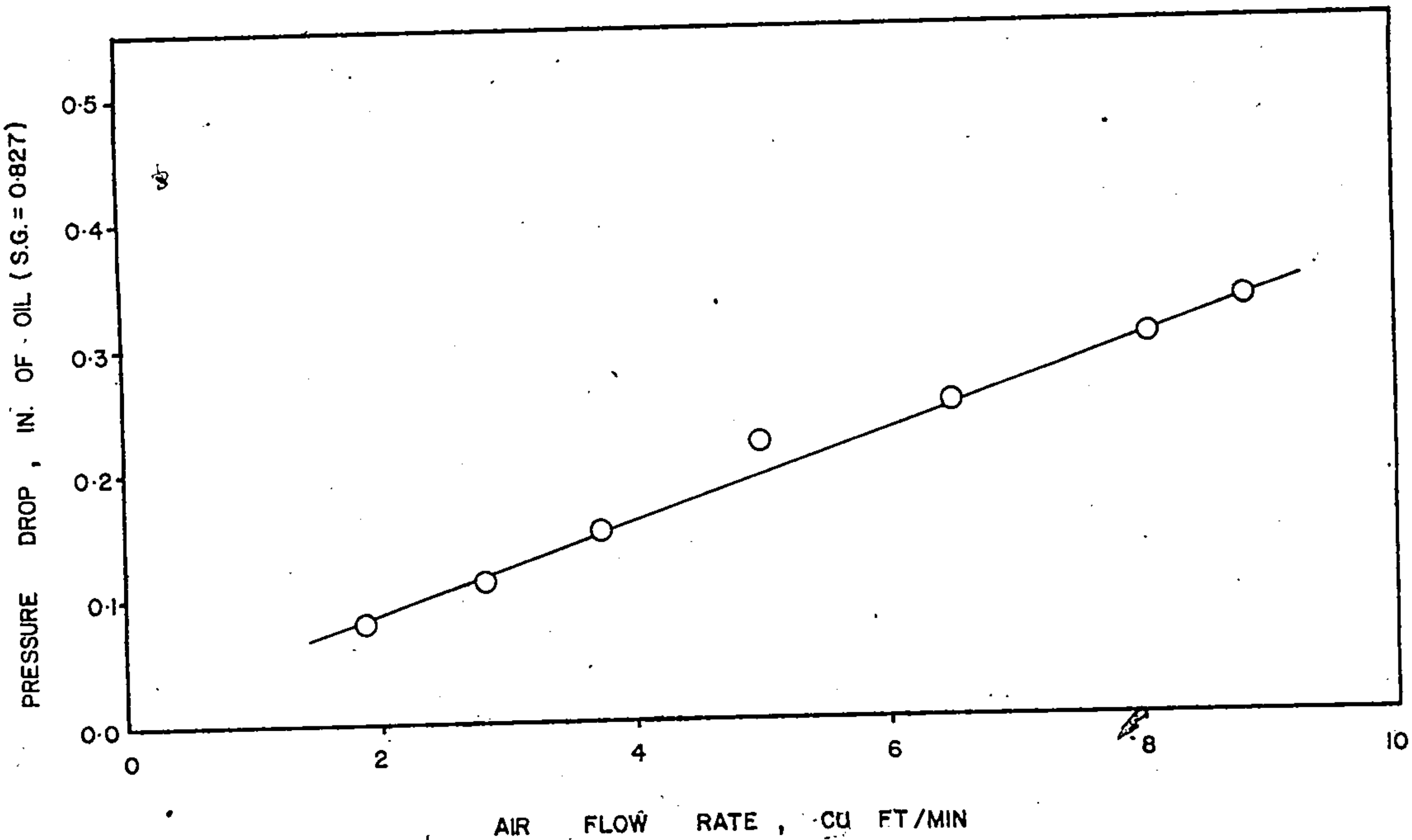


Fig. 17

Pressure drop in empty column

Design of Cyclone Separator[10]

Inlet velocity to cyclone is the most critical parameter deciding smallest particle size that will be collected. As the velocity is reduced, particle size settling increases. Hence, for design, minimum possible flow of air is considered. The design basis selected is given below.

Inlet cross section = $1.0 \times 2.0 \text{ cm}^2$

Air velocity = 100 cm/min

Other dimensions of the cyclone are selected as recommended

Calculations

Minimum air flow = velocity x bed area

$$= 100 \times \pi \times \frac{(24.94)^2}{4.0} = 48827.3 \text{ cm}^3/\text{min}$$

$$\text{Inlet air velocity} = \frac{48827.3}{2.0 \times 60.0} = 406.9 \text{ cm/sec}$$

Min. particle diameter collected is given as :

$$D_P^2 \min = \frac{9 \mu B_c}{5 \pi V_c (\rho_s - \rho)}$$

D_P min	-	min. particle diameter
μ	-	air viscosity
B_c	-	width of air inlet
V_c	-	inlet velocity
ρ_s	-	solid density
ρ	-	air density

$$D_P^2 = \frac{9 \times 0.018 \times 10^{-2} \times 1.0}{5\pi \times 406.9 \times 1.5} \qquad \rho_s \gg \rho$$

$$= 16.9 \times 10^{-8} \text{ cm}^2$$

Therefore $D_{P_{min}} = 4.1 \times 10^{-4}$ cm

$= 4.1\ \mu$

which is quite sufficient.

Collection efficiency for particles of $23\ \mu$

$$D_{PC} = \frac{D_P}{\sqrt{2}} = \frac{4.1}{\sqrt{2}} = \frac{4.1}{1.414} = 2.9\ \mu$$

$$D_P = 23\ \mu$$

$$\frac{D_P}{D_{Pc}} = \frac{23}{2.9} = 7.59$$

from the graph(10) collection efficiency = 98%

Pressure drop in cyclone

$$\Delta P = k \left(\frac{D_c}{D_e} \right)^2$$

where

ΔP - in. of water

D_c, D_e - dimensions of cyclone in fig. 18

$$\Delta P = 3.2 \left(\frac{4.0}{2.0} \right)^2$$

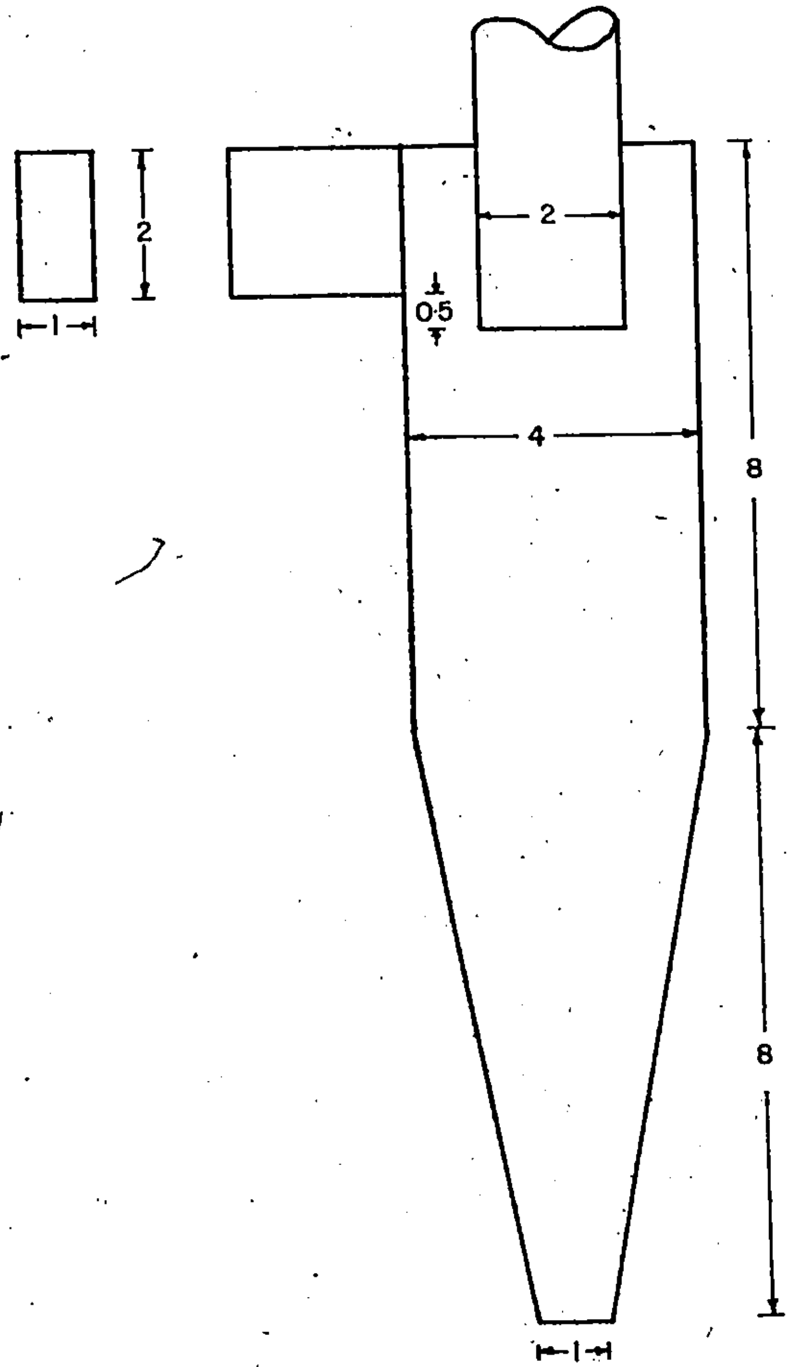
$$= 3.2 \times 4 = 12.8 \text{ in. of water}$$

Dimensions of screw feeder

Total length	= 55 cm
Diameter	= 1.7 cm
Pitch	= 3.0 cm
Static torque requirement	= 354 gm/cm

Dimensions of output screw conveyor

Total length	= 39 cm
Diameter	= 1.7 cm
Pitch	= 2.7 cm
Static torque requirement	= 393 gm/cm



ALL DIMENSIONS IN CM

Fig. 18

Cyclone Separator

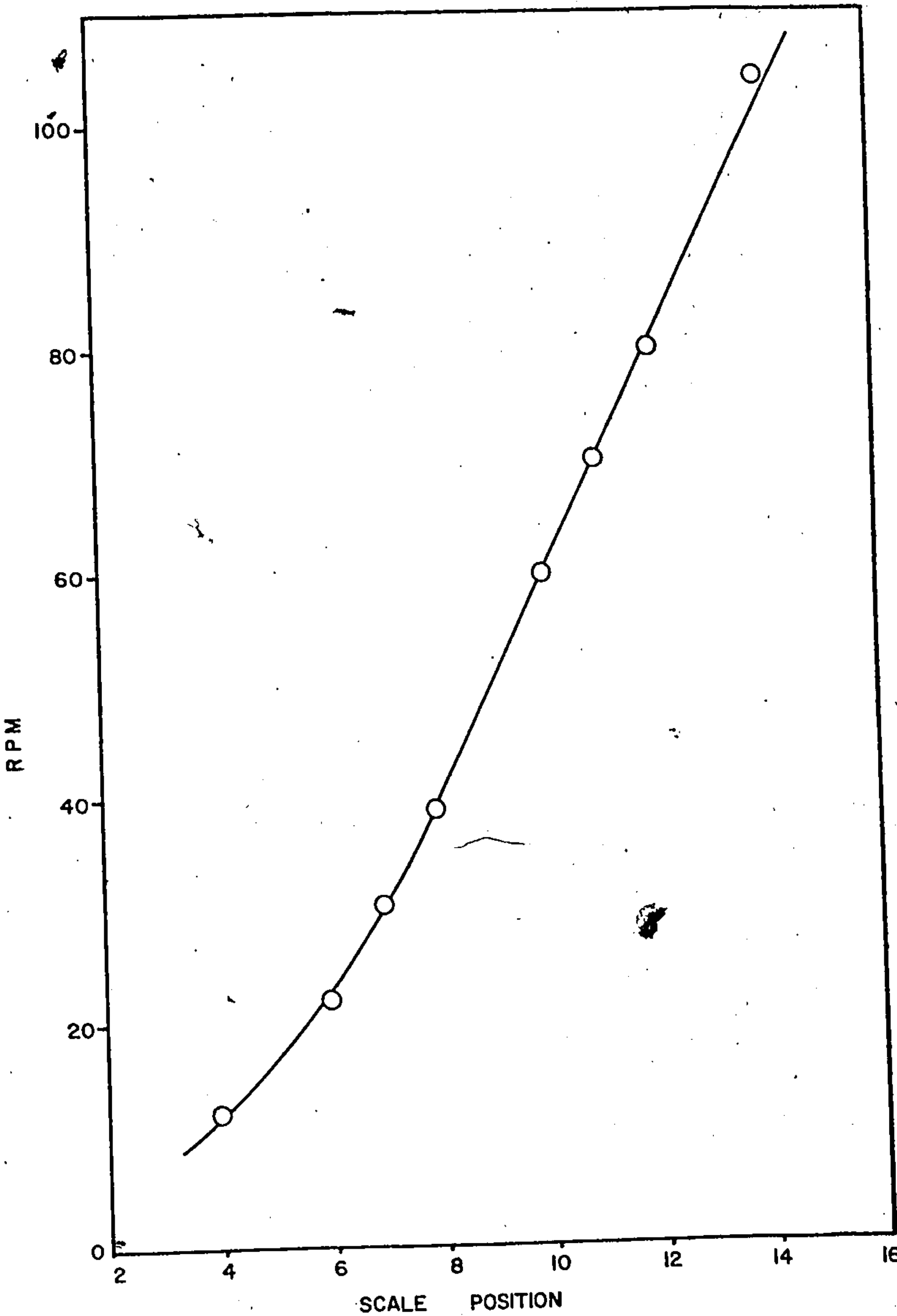


Fig. 19(a)

RPM vs. Scale Position

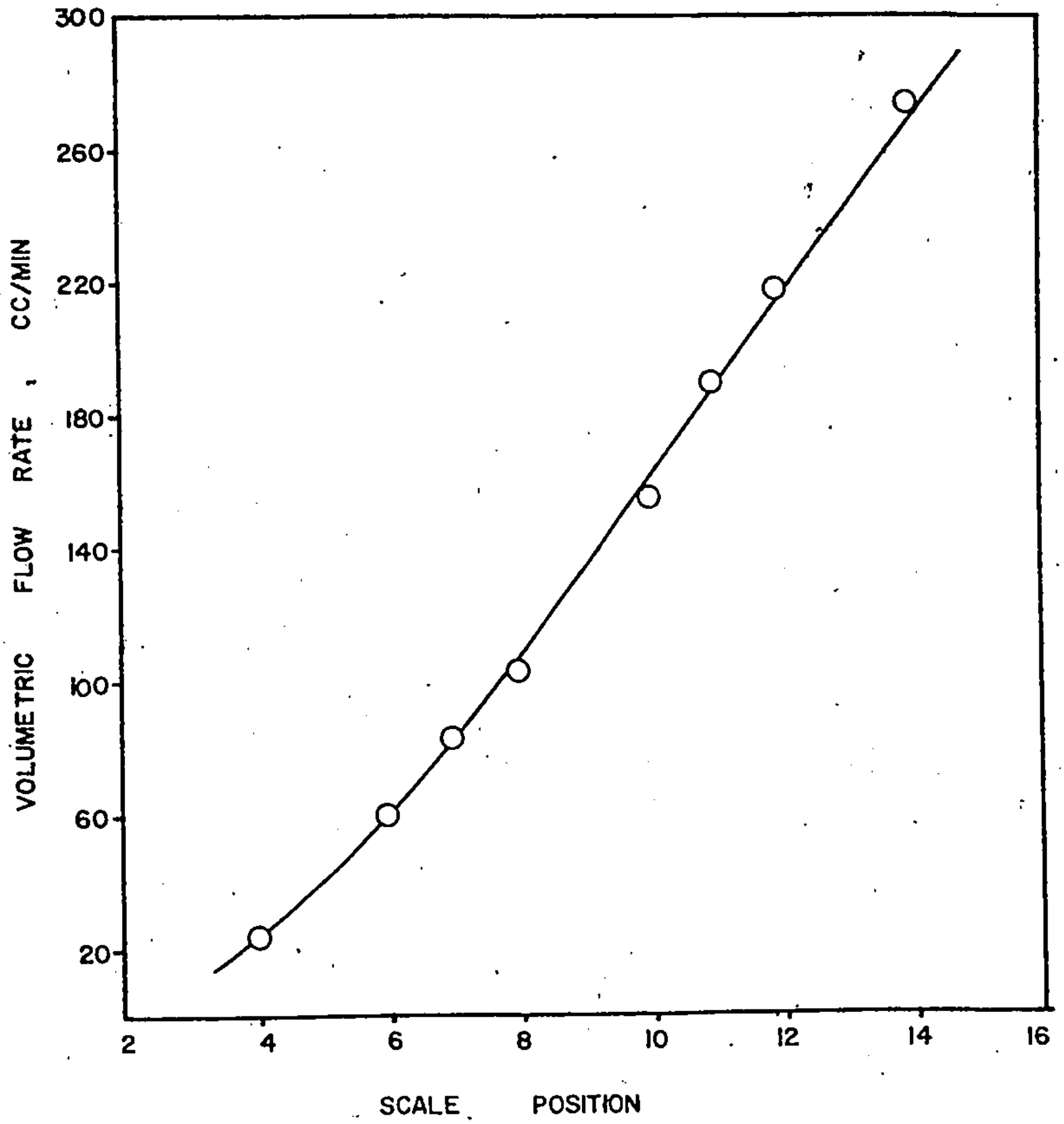


Fig. 19(b)

Volumetric flow rate vs. Scale Position

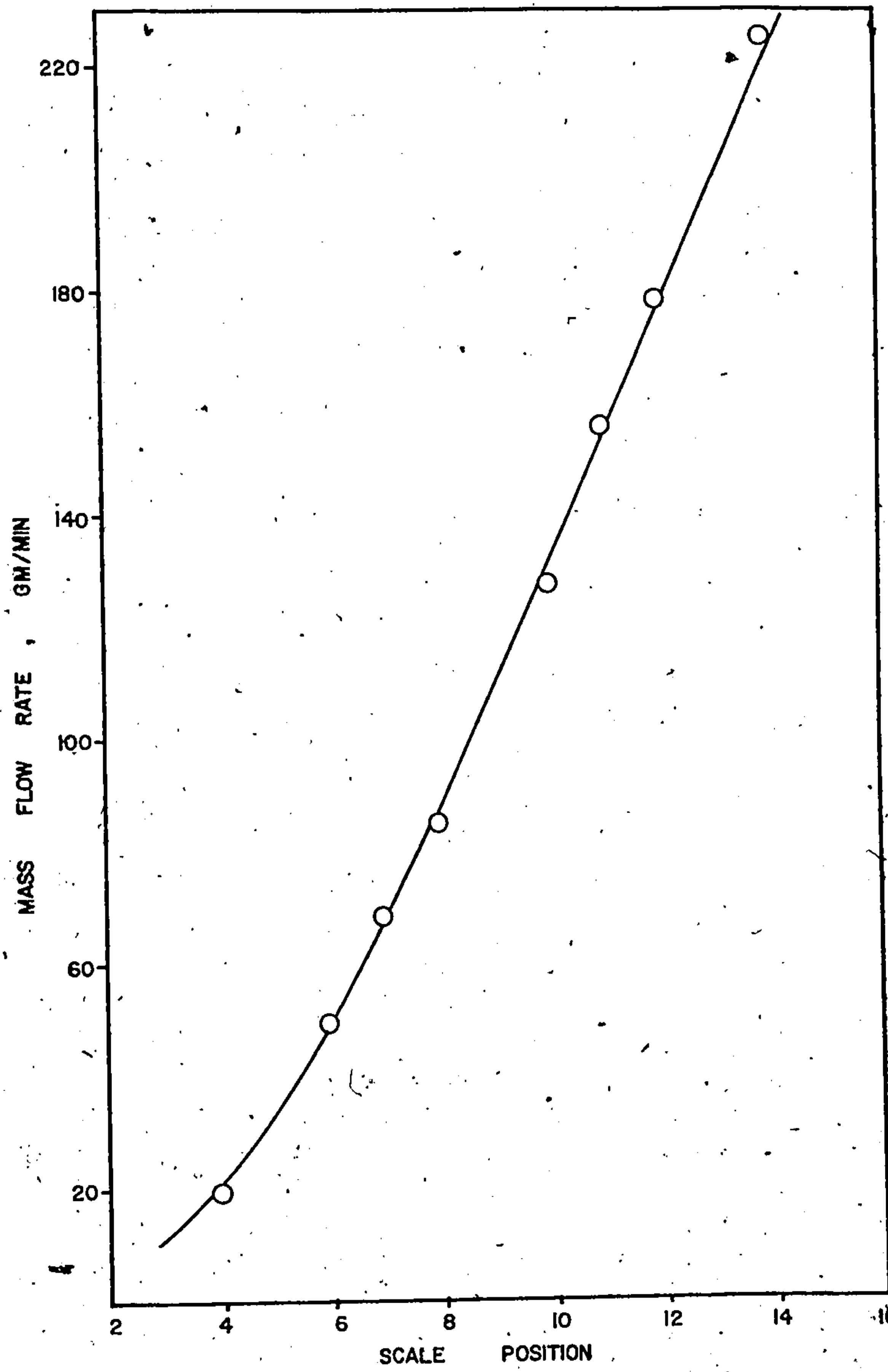
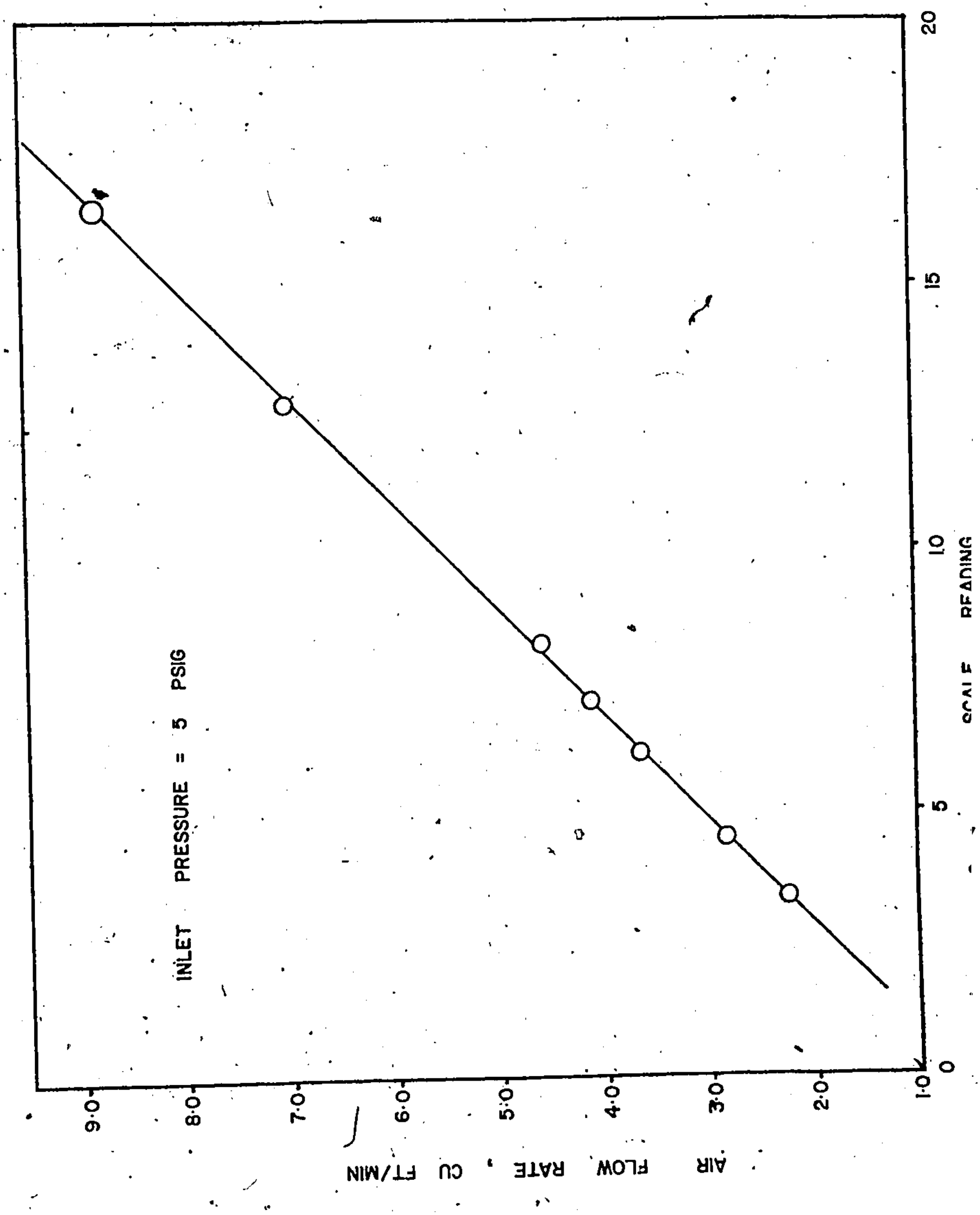


Fig. 19(c)

Mass flow rate vs. Scale Position


INLET PRESSURE = 5 PSIG
AIR FLOW RATE , CU FT/MIN
1·0
2·0
3·0
4·0
5·0
6·0
7·0
8·0
9·0
0
5
10
15
20