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PROGRESS REPORT

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Forward

This report summarizes the major findings and results of the various studies carried out under Contract DE-AC02-79-ER-10386, "Particle Deposition in Granular Media" during the period from February 1, 1986 to date. The report consists of six parts each of which is devoted to a separate topic of investigation. For convenience, the equations, figures and tables of each part are enumerated separately. Separate sections on references and notations are provided for each part.

Part I.

Experimental Determination of the Initial Collection

Efficiency in Granular Beds - An Assessment of

Effect of the Instrument Sensitivity and the

Extent of Particle Bounce-off

Introduction

A granular bed's capability for collecting particles from gas streams passing through the bed can be assessed in terms of the overall collection efficiency, E , or, alternatively, the penetration, P , defined as

$$E = 1 - P = 1 - \frac{c_{eff}}{c_{in}} \quad (1)$$

where c_{eff} and c_{in} denote, respectively, the effluent and influent particle concentrations.

The overall collection efficiency, E , (or the penetration, P) is a function of a number of variables, such as bed height, filter-grain size, gas velocity, etc. A more fundamental quantity which represents the granular bed's intrinsic ability to capture particles is the so-called single-collector (or unit-collector) efficiency, η . Single-collector efficiency is a function of the size of the filter grains, the size of the particles to be captured, the flow conditions around the filter grains, and the extent of deposition on the filter grains. Initially, when extent of deposition is slight and the bed is homogeneously packed, the single-collector efficiency throughout the bed is constant (or $\eta = \eta_0$). η_0 and E are related by the expression

$$E = 1 - \frac{c_{eff}}{c_{in}} = 1 - \exp \left[K_1 (1 - \epsilon) \frac{L}{d_g} \eta_0 \right] \quad (2)$$

where L is the height of the filter; ϵ , its porosity; and d_g , the filter grain diameter. K_1 is constant and is assumed to be 1.5 (Snaddon and Dietz, 1980; D'Ottavio and Goren, 1983), $\frac{1.5}{\epsilon}$ (Paretsky et al., 1971; Tardos et al., 1974), or $\left[\frac{6}{(1 - \epsilon)^{2/3}} \right]^{1/3}$ (Tien and Payatakes, 1979; Pendse and Tien, 1980, Yoshida and Tien, 1985).

Equation (2) provides the basis for determining η_0 from experiments. By conducting measurements using experimental filters, one can readily determine the effluent and influent concentrations, which together with the grain diameter, filter height, and bed porosity, can be used to calculate η_0 from the total efficiency data according to Equation (2). The multiple values available for K_1 do not present any real problem since η_0 obtained by using one value of K_1 can be readily converted to those based on the other values. It is, however, an inconvenience when values of η_0 obtained from different sources are compared.

An important factor affecting the accuracy of determining η_0 experimentally derives from the fact that η_0 is proportional to the logarithm of the concentration ratio ($\ln \frac{c_{in}}{c_{eff}}$). Generally speaking, particle concentration in gas streams is a quantity which cannot be measured easily or very accurately. A relatively modest inaccuracy in c_{in} or c_{eff} often results in a highly significant error in η_0 , as shown in the hypothetical error propagation analysis by Pendse and Tien (1980).

In recent years a number of investigators have reported results on collection efficiency in granular beds: Thambimuthu, 1980; D'Ottavio and Goren, 1983; Gal et al., 1985; Walata et al., 1986; Takahashi et al., 1986. Without exception, these investigators used commercially available particle counters to measure concentration. While using these particle counters represents a significant improvement over earlier methods for determining particle concentration (cascade impactor, for example), counters are prone to error. For example, Yokochi (1985) observed that several different types of particle counters produced substantially different results in particle concentration under identical conditions. Furthermore, a given type of particle counter offered significantly different results, depending upon its length of service.

The purpose of the present study is three-fold. The first part of the study was devoted to determining the particle collection efficiency of clean granular beds by using three different types of particle counters. The results obtained were examined and compared to determine the possible instrument dependency. Second, the substantial amount of new data collected was used as a basis for assessing the relative accuracy of the various empirical correlations of η_0 suggested by investigators in recent years. Finally, based on the data collected, new correlations for η_0 and the adhesion probability γ were developed. The experiments were conducted under the inertial-impaction-dominated regime. For aerosol filtration, inertial impaction is considered one of the most important and often the dominant mode of particle collection.

Experimental Work

Apparatus. A schematic diagram of the apparatus used for the experimental work is shown in Fig. 1. The apparatus is essentially the same as that used in the past (Walata et al., 1986) and is composed of four sections: the aerosol generating section, the heating section, the neutralizing section, and the test section.

A collision atomizer (BGI, Inc., Waltham, MA) was used to generate test aerosols of monodispersed latex particles. Bottled nitrogen gas (at elevated pressure) was passed through the aspirators of the atomizer to create a spray. The spray then entered the heating section, which was made of a glass tube (5/8 in. i.d.) wrapped with heating tape. The function of the heating section was to vaporize any water adhering to the latex particles.

The aerosol then passed through a neutralizer (TSI, Inc., Model 3012, St. Paul, MN) in order to reduce the electrostatic charges on the latex particles and finally the test section.

The test section consisted of an experimental filter preceded by an aerosol tunnel. The experimental filter was made of a plexiglass tube (5 in. in height and 1.5 in. in diameter). A recess was milled into the bottom of the tube so that a support screen and a lock ring could be inserted. The aerosol tunnel, which consisted of an expanding section, a section of constant cross-sectional area, and a contracting section, was constructed from a layer of epoxy resin and fiberglass on a smooth, waxed wooden pattern. The purpose of the aerosol tunnel was to insure a uniform flow field as the test aerosols entered the experimental filter.

The test section included two aerosol sample probes, the first in the upper part of the aerosol tunnel, and the second immediately after the filter. A pressure tap was installed on each end of the filter.

Particle Concentration Determinations. The aerosol concentrations of both the influent and effluent samples were determined using three different particle counters: a Climet Particle Analyzer (Model 208, Climet Instrument Co., Redland, CA); an Aerodynamic Particle Sizer (APS 33 Aerodynamic Particle Sizer, TSI, Inc., St. Paul, MN); and a HIAC/Royco Counter (Model PA-720, HIAC/Royco, CA). Both the Climet and the HIAC/Royco counters operate on the principle of optical blockage; the Aerodynamic Particle Sizer was a laser velocimeter to determine particle size by measuring the particle's accelerating speed. Although the capabilities of these three counters are not identical, they may be considered

comparable in their ability to determine dilute particle concentrations with particle diameter of $1 \sim 10 \mu\text{m}$. All three counters have been used in the previous studies of collector efficiency of granular filtration.

Experimental Systems. Glass beads within a narrow size range were used as filter grains. Microscopic examination of the beads indicated that they were spherical, with smooth surfaces. The bed porosity was determined for each measurement and was constant (varying from 0.33 to 0.35).

The test aerosols were prepared from aqueous suspensions of polyvinyl-toluene latex particles supplied by the Dow Chemical Co. (Indianapolis, IN).

Experimental Variations. The experimental variables examined were the types of particle counters used (Climet Counter Model 208, TSI APS 33 Aerodynamic Sizer; and HIAC/Royco Model PA-720 Counter); filter-grain size ($108 \mu\text{m}$, $244 \mu\text{m}$, and $508 \mu\text{m}$ diameter); aerosol size ($2.02 \mu\text{m}$ and $4.1 \mu\text{m}$ in diameter); bed height (0.142 cm , 0.34 cm , and 0.656 cm); and gas flow rate ($u_g = 5.55, 11.10, 27.74, 55.48 \text{ cm/s}$).

Results

Many measurements were taken. Each measurement lasted approximately one hour, during which samples of influent and effluent were analyzed for their particle concentrations. In addition, pressure drop changes were recorded even though in all cases these changes occurring during an experimental run were always negligible.

Accuracy of the Experimentally Determined η_0 . The experimental accuracy of the initial collection efficiency can be assessed by considering the following: the internal consistency of η_0 obtained from a given experimental run; the reproducibility of η_0 obtained from independent runs carried out under similar conditions and using the same particle counter; and, finally, comparison of

the results obtained under similar conditions but with different particle counters.

The internal consistency of the data can be seen by comparing the η_o values obtained at different times in an experimental run. Note that since very dilute aerosol suspensions were used, the extent of particle collection during an experimental run of approximately one hour was slight. Thus, the effect of particle deposition can be ignored. In principle, the values of η_o obtained at different times should remain the same.

η_o was determined from experimental data (in the form of c_{in} and c_{eff} vs. time according to Equation (2) which upon rearrangement gives

$$\eta_o = - \frac{dg}{L} \frac{1}{K_1(1-\epsilon)} \ln \frac{c_{eff}}{c_{in}} \quad (3)$$

where K_1 is given as

$$K_1 = \left[\frac{6}{(1-\epsilon)^{2/3}} \right]^{1/3} \quad (4)$$

The experimental results of c_{in} vs. time, c_{eff} vs. time, and (c_{eff}/c_{in}) vs. time for three independent runs using the Climet counter are shown in Figs. 2a and 2b. The three runs whose results appear were conducted under similar conditions except for influent concentrations.

In all three cases, the influent concentration for a given run was found to fluctuate somewhat. This behavior reflects at least partially the inherent limitations of the collision atomizer used in generating test suspensions at constant concentrations. It may also reflect the inaccuracy of the particle counter used. The same fluctuating behavior was observed in the effluent concentration history, but the extent of fluctuation was much less.

The ratio $c_{\text{eff}}/c_{\text{in}}$ (calculated from the concentration histories against time) is shown for the three runs in Fig. 2b. Although the influent concentration levels of these three runs were different, the concentration ratio remained essentially the same. This result is consistent with the expectation (as well as the theories) that during the initial filtration period, and for the inertial-impaction-dominated regime, the extent of particle collection is independent of the influent concentration. Fig. 2b also shows that both the internal consistency of the result obtained from a given run and the reproducibility of the experimental data are good. Similar conclusions can be drawn for experiments that use other types of particle counters. For example, see Figs. 3a and 3b for similar results obtained using a HIAC/Royco counter.

The single collector efficiency, η_0 , obtained from the results shown in Fig. 2b is given in Fig. 4. Here the η_0 obtained from a given run as well as those derived from varied runs differ more widely than did the concentration ratios depicted in Fig. 2b. This difference in the extent of scattering, is, of course, a direct consequence of the logarithmic relationship between the concentration ratio and η_0 given by Equation (2). Similar behavior was also observed for the data obtained using the HIAC/Royco and TSI counters [see Fig. 5].

The results shown in Figs. 3 and 4 indicate that the variation in η_0 within a given run is comparable to the variations in η_0 obtained from different runs. This behavior can be seen more clearly from the numerical results given in Table 1, which summarizes the average values of η_0 for each experimental run and the variations in results in terms of the ratio of the standard deviation to the average value of η_0 , $\bar{\sigma}_{\eta_0}/\bar{\eta}_0$, for the six runs whose results are shown in Figs. 4 and 5. It is clear that the average values of $\bar{\eta}$ obtained from

independent runs but under identical operating conditions (except the influent concentration) and using the same particle counter remain remarkably constant, and the ratio $\sigma_{\bar{\eta}_0} / \bar{\eta}_0$ remains quite similar too. These results demonstrate that the experiments yielded both consistency and good reproducibility.

The experimental results were analyzed to determine whether or not using different types of particle counters contributes to a difference in the values for η_0 or whether or not there is a relationship between the type of particle counter used and the internal consistency of the data obtained. The analysis proceeded as follows. As shown above, each of the three types of particle counters gave good reproducibility of results. The results obtained from independent runs carried out under identical conditions (namely, same superficial velocity, aerosol-particle size, and filter-grain size) and using the same type of particle counter were pooled. From these results the mean value of η_0 and the standard deviation were calculated, as shown in Table 2. [For example, all the results of Run Nos. 860301, 860304, and 860306 are now combined into a single entry in Table 2.]

The results summarized in Table 2 lead to two conclusions. First, in spite of the variation in η_0 observed before, the average values of η_0 corresponding to the three different types of particle counters are very close. Even in the worst case, the difference is not more than 40%. Moreover, there is no evidence to suggest any bias in the values of η_0 that could be attributed to using a particular type of counter. In the second to the last column of Table 2, the types of particle counters are ranked by η_0 value in descending order (with the letters C, H, and T denoting the Climet Counter, HIAC/Royco Counter, and TSI Particle Sizer, respectively). It is obvious that the ranking does not follow any particular pattern.

The second conclusion is also rather obvious, namely, the degree of scattering of $\bar{\eta}_0$ corresponding to a fixed set of operating conditions (namely, u_g , d_p , and d_g) varies with the type of the particle counter used. In the last column of Table 2, the particle counters are ranked according to their values $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$. The TSI Aerodynamic Particle Sizer values were found to vary the most; the HIAC/Royco counter the least.

It is natural to inquire whether the extent of variation is related to the concentration of the test suspensions used in the experiments. For this purpose, the values of $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ are plotted against the average influent concentration, $\bar{c}_{in}$, for each experimental run and are shown in Figs. 6a, 6b, and 6c (corresponding to the three types of particle counters used). Although the data points in all three cases display considerable scattering, it is clear that on the average, the HIAC/Royco counter gives the best internal consistency for η_0 (or in other words, the smallest value of $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$).

Some evidence seems to indicate that this difference in the values of $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ can be related to the characteristics of the particle counters involved. As shown in Figs. 2a and 3a, the influent concentrations for a given experimental run were not exactly constant. The fluctuation in c_{in} can be expressed as $\sigma_{c_{in}}/\bar{c}_{in}$, where $\bar{c}_{in}$ is the mean value of c_{in} and $\sigma_{c_{in}}$ the standard deviation. In Figs. 7a, 7b, and 7c, the values of $\sigma_{c_{in}}/\bar{c}_{in}$ for the three counters are plotted vs. $\bar{c}_{in}$. By comparing Figs. 7a, 7b, and 7c against Figs. 6a, 6b, and 6c, it becomes obvious that the magnitudes of the two quantities $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ and $\sigma_{c_{in}}/\bar{c}_{in}$ are comparable. Furthermore, the relationship of $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ vs. $\bar{c}_{in}$ is also similar to that of $\sigma_{c_{in}}/\bar{c}_{in}$ vs. $\bar{c}_{in}$. For example, both Figs. 6c and 7c exhibit a decrease in the ratio of the standard deviation to the mean value ($\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ or $\sigma_{c_{in}}/\bar{c}_{in}$) as $\bar{c}_{in}$ increases. Similarly, this decreasing behavior is not evident either in Fig. 6a or 7a.

Comparisons with Empirical Correlations. Over the years, a number of investigators have proposed correlations for single-collector efficiency of granular beds. A summary of these correlations is listed in Table 3. Also included in this table is the functional relationship between the total collection efficiency, E , and the single-collector efficiency, η_o , which were used to obtain η_o from experimental data.

A casual examination would reveal the significant differences among these empirical correlations. Pendse and Tien (1980) showed that some of these earlier correlations yield results which may differ by order of magnitude. Even among the more recent correlations, some significant differences exist. For example, the correlation developed by D'Ottavio and Goren (1983) is based on a modified Stokes number which is a function of the standard Stokes number, the Reynolds number, and the bed porosity. On the other hand, Thambimuthu's correlation ignores the possible effect of the Reynolds number because the data on which that correlation is based does "not permit the Stokes and Reynolds numbers to be distinguished as independent variables" (Clift, 1983).

The more recent experimental work of Gal et al. (1985), however, shows that quite different single-collector efficiencies were obtained using air and helium as carrier gases at identical conditions. Since the viscosities of helium and air are similar but the density of helium is only one-seventh that of air, the observed differences in the single-collector efficiency can only be explained through the Reynolds number. This information casts some uncertainty on the validity of ignoring the Reynolds number in developing empirical correlations of η_o .

In spite of the uncertainties, however, it is obvious that the more recently established correlations offer considerable improvement over the earlier ones. The chief advantages are that the data on which these newer correlations are based are more extensive and are likely to be more accurate because of recent advanced particle counting instrumentation. Thus, the experimental results obtained were only compared with the four more recent correlations, namely, those of Thambimuthu (1980), D'Ottavio and Goren (1983), Gal et al. (1985), and Yoshida and Tien (1985).

Correlation by Thambimuthu

$$\eta_o = \left[\frac{N_{St}}{N_{St} + 0.062} \right]^3 \quad (5)$$

$$\text{for } 1.26 \leq N_{Re} \leq 130$$

$$0.33 \leq \quad \leq 0.4$$

$$2 \times 10^{-3} \leq N_{St} \leq 2 \times 10^{-2}$$

where

$$N_{Re} = d_g \rho u / \mu \quad (6)$$

$$N_{St} = \rho_p d_p^2 u c / 9 \mu d_g \quad (7)$$

Correlation by D'Ottavio and Goren

$$\eta_o = \frac{(N_{St_{eff}})^{3.35}}{1.67 + (N_{St_{eff}})^{3.35}} \quad (8)$$

where

$$N_{St_{eff}} = [A(\alpha) + 1.14 N_{Re}^{1/2} (1-\alpha)^{-3/2}] (N_{St}/2) \quad (9)$$

$$A(\alpha) = \frac{6 - 6\alpha^{5/3}}{6 - 9\alpha^{1/3} + 9\alpha^{5/3} - 6\alpha^2} \quad (10)$$

$$\alpha = 1 - \epsilon \quad (11)$$

Correlation by Gal et al.

$$\eta_o = \frac{2 N_{St}'^{3.9}}{4.3 \times 10^{-4} + N_{St}'^{3.9}} \quad (12)$$

where

$$N_{St}' = N_{St} \cdot F \quad (13)$$

$$F = 1.0 + 1.75 N_{Re} / [150(1-\epsilon)] \quad (14)$$

Correlation by Yoshida and Tien

$$\eta_o = B [100 N_{St}^2 + 0.19 (4 - \frac{4N_R}{d_c^*} - \frac{N_R^2}{d_c^{*2}})^{1/2} \frac{N_R^{1.041}}{d_c^*}] \quad (15)$$

$$\text{for } N_R < 0.002$$

$$N_{St} < 0.01$$

$$\eta_o = B [N_{St} + 0.48 (4 - \frac{4N_R}{d_c^*} - \frac{N_R^2}{d_c^{*2}})^{1/2} \frac{N_R^{1.041}}{d_c^*}] \quad (16)$$

$$\text{for } N_R > 0.002$$

$$N_{St} < 0.01$$

$$B = 7 - 6 \exp [-0.0065 N_{Re}] \quad (17)$$

where d_c^* is the dimensionless constriction diameter, defined as d_c/d_g . For a bed composed of nearly mono-sized filter grains, d_c^* may be taken to be 0.35.

It should be noted that the first three correlations were developed by the respective investigators based on their own experimental data. The correlation of Yoshida and Tien is a modified version of the earlier correlation proposed by Pendse and Tien (1980) and based on data reported by a number of investigators, including D'Ottavio and Goren and Thambimuthu. Furthermore, the single-collector efficiency in the Yoshida and Tien correlation was based on $K_1 = \left[\frac{6}{(1-\epsilon)^{2/3}} \right]^{1/3}$, while for the other three correlations, η_o is based on $K_1 = 3/2$.

The current experimental data were compared with the four correlations, and the results are summarized in Table 4. The degree of agreement can be seen from the ratio $(\eta_o)_{\text{pred}}/(\eta_o)_{\text{exp}}$, where the subscripts "pred" and "exp" denote, respectively, the predicted and experimental values. The results indicate that the Yoshida and Tien correlation agrees most closely with the experimental data, followed by the correlation of Thambimuthu. The correlations of both D'Ottavio and Goren and Gal et al. tend to severely underestimate η_o at low Stokes numbers.

The fact that the experimental results failed to agree with the correlation of D'Ottavio and Goren requires some consideration since that correlation was developed from an extensive body of experimental data. In Table 5 are listed the experimental conditions used by D'Ottavio and Goren in their work and those under which the present study was conducted. While many of the experimental variables used in these two studies are comparable, the most important difference is that the relative size parameters (d_p/d_g , or the interception parameter) in D'Ottavio and Goren's work was much smaller than that used in this study. Their correlation also appears to underestimate their own data (for example, see Figs. 9 and 10 of D'Ottavio and Goren, 1983). These two factors appear responsible for the underestimation mentioned earlier.

To further substantiate this assessment, the experimental results obtained in this work as well as those reported by D'Ottavio and Goren were plotted against the parameter $N_{St_{eff}}$, as shown in Fig. 8. It is obvious that the data points fall into two distinct groups. For a given $N_{St_{eff}}$, the data obtained in this study (and consequently higher values of N_R) are generally greater than those of D'Ottavio and Goren. Thus, it appears that η_o cannot be considered as a function of $N_{St_{eff}}$ alone; the effect of the relative size parameter must also be included.

New Correlation of η_o . A simple way of incorporating the effect of N_R is to express η_o/N_R^a as a function of $N_{St_{eff}}$. By trial and error, it was found that the best result was obtained if "a" is taken to be 0.23 (see Fig. 9). The following expression is obtained from the data using glass beads as collector:

$$\eta_o = 0.2589 N_{St_{eff}}^{1.3437} \cdot N_R^{0.23} \quad \text{for } N_{St} < 1.2 \quad (18)$$

$$1.5 \times 10^{-4} < N_R < 4.5 \times 10^{-2}$$

For $N_{St_{eff}} > 1.2$, η_o is shown to decrease with the increase in $N_{St_{eff}}$, because the aerosol particles tend to bounce off the filter grains upon impact with them. The single-collector efficiency can now be expressed as

$$\eta_o = 0.2589 \cdot N_{St_{eff}}^{1.3437} \cdot N_R^{0.23} \cdot \gamma \quad (19)$$

where γ is the so-called adhesion probability. The adhesion probability can be estimated from experimental data by dividing the experimentally determined value of η_o by the value calculated from Equation (18). A plot of γ vs.

$N_{St_{eff}}$ is shown in Fig. 10.

The adhesion probability can be expressed as

$$\gamma = 1.4315 N_{St_{eff}}^{-1.968} \quad (20)$$

Conclusions

The initial collection efficiency in granular beds was determined experimentally to examine the possible effect of the particle counters used in the experiments. It was found that depending upon the particular type of counter used to determine aerosol concentrations, the value of η_0 exhibits considerable scattering but the mean values remain markedly constant. The experimental data were also used to develop a new empirical correlation, taking into account the effect of the parameter $N_R = d_p/d_g$.

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Table 1

Internal Consistency of Experimentally Determined η_o

Run No.	$\bar{\eta}_o^*$	$\sigma_{\bar{\eta}_o}$	$\sigma_{\bar{\eta}_o}/\bar{\eta}_o$	$\sigma_{c_{in}}^-/\bar{c}_{in}$
860301 ⁺	6.6×10^{-3}	2.36×10^{-3}	0.358	0.098
860304 ⁺	1.0×10^{-2}	3.6×10^{-3}	0.360	0.058
860306 ⁺	1.0×10^{-2}	1.95×10^{-3}	0.195	0.026
860302 [†]	6.17	2.22×10^{-3}	0.36	0.086
860303 [†]	1.06×10^{-2}	1.98×10^{-3}	0.18	0.074
860305 [†]	7.7×10^{-3}	8.47×10^{-4}	0.11	0.050

⁺Climate Counter used in experiment[†]H/R Counter used in experiment

*Based on six data points taken over a period of approximately 1 hr. The experimental conditions are:

$$d_p = 2.02 \mu\text{m}$$

$$d_g = 508 \mu\text{m}$$

$$L = 0.69 \text{ cm}$$

$$\epsilon = 0.32 \sim 0.33$$

$$u_s = 5.5 \text{ cm/s}$$

Table 2. Summary of Experimental Results of Initial Collection
Efficiency Obtained Using Three Different Particle Counters

Experimental Condition			H/R Counter			TSI APS Counter			Climet Counter			Ranking of Particle Counter								
(cm/sec)	dp(μm)	dg(μm)	$\bar{c}_{in}$ (H/R)	$\bar{c}_{in}$ (TSI)	$\bar{c}_{in}$ (Climet)	$\bar{n}_P$	σ_P	$\sigma_P/\bar{n}_P$	$\bar{n}_T$	σ_T	$\sigma_T/\bar{n}_T$	$\bar{n}_c$	σ_c	$\sigma_c/\bar{n}_c$	According to $\bar{c}_{in}$		According to $\sigma_{in}/\bar{n}_0$			
5.55	2.02	508	56.25		48.29	8.16×10^{-3}	2.49×10^{-3}	0.305				8.93×10^{-3}	3.07×10^{-3}	0.343	C	H	C	H		
11.10	"	"	19.26	23.29	14.84	5.82×10^{-3}	4.26×10^{-4}	0.073	9.84×10^{-3}	5.59×10^{-3}	0.568	7.85×10^{-3}	2.79×10^{-3}	0.355	T	C	H	T	C	H
27.74	"	"	8.05	5.07	3.75	3.20×10^{-2}	9.83×10^{-3}	0.307	3.00×10^{-2}	1.69×10^{-2}	0.563	3.01×10^{-2}	5.30×10^{-3}	0.176	H	C	T	T	H	C
55.48	"	"	5.23	3.97	2.47	7.25×10^{-2}	1.04×10^{-2}	0.143	8.86×10^{-2}	3.43×10^{-2}	0.387	7.53×10^{-2}	1.36×10^{-2}	0.181	T	C	H	T	C	H
5.55	2.02	244	178.46		103.44	1.42×10^{-2}	5.65×10^{-3}	0.398				1.61×10^{-2}	3.54×10^{-3}	0.220	C	H		H	C	
11.10	"	"	73.87	87.25	61.39	2.74×10^{-2}	3.96×10^{-3}	0.144	2.60×10^{-2}	8.64×10^{-3}	0.332	2.53×10^{-2}	4.27×10^{-3}	0.169	H	T	C	T	C	H
27.74	"	"	28.62	25.87	23.75	1.82×10^{-1}	1.73×10^{-2}	0.095	1.71×10^{-1}	2.53×10^{-2}	0.148	1.76×10^{-1}	2.12×10^{-2}	0.120	H	C	T	T	C	H
27.74	"	"	7.47	9.28	7.00	1.52×10^{-1}	1.69×10^{-2}	0.111	1.46×10^{-1}	3.11×10^{-2}	0.213	1.41×10^{-1}	1.73×10^{-2}	0.123	H	T	C	T	C	H
55.48	"	"	4.43	3.86	3.34	1.27×10^{-1}	1.38×10^{-2}	0.109	1.12×10^{-1}	3.35×10^{-2}	0.299	1.21×10^{-1}	2.00×10^{-2}	0.165	H	C	T	T	C	H
5.55	4.1	244	10.82		4.67	1.34×10^{-1}	1.11×10^{-2}	0.083				1.43×10^{-1}	3.04×10^{-2}	0.213	C	H		C	H	
11.10	"	"	6.28	6.63	3.13	2.03×10^{-1}	2.52×10^{-2}	0.124	1.86×10^{-1}	2.77×10^{-2}	0.149	2.12×10^{-1}	2.50×10^{-2}	0.118	C	H	T	T	H	C
27.74	"	"	2.63	2.00	1.41	7.45×10^{-2}	9.30×10^{-3}	0.125	7.70×10^{-2}	3.20×10^{-2}	0.416	6.11×10^{-2}	1.15×10^{-2}	0.188	T	H	C	T	C	H
55.48	"	"	0.53	0.40	0.32	2.53×10^{-2}	7.35×10^{-3}	0.291	2.07×10^{-2}	1.13×10^{-2}	0.546	3.22×10^{-2}	7.42×10^{-3}	0.230	C	H	T	T	H	C
5.55	2.02	103	45.15		42.09	7.20×10^{-2}	7.94×10^{-3}	0.110				7.96×10^{-2}	7.85×10^{-3}	0.099	C	H		H	C	
11.10	"	"	20.36	20.67	16.78	1.27×10^{-1}	2.28×10^{-2}	0.180	1.36×10^{-1}	2.67×10^{-2}	0.196	1.62×10^{-1}	1.17×10^{-2}	0.072	C	T	H	T	H	C
27.74	"	"	7.43	7.50	6.43	7.24×10^{-2}	1.93×10^{-3}	0.027	9.89×10^{-2}	1.09×10^{-2}	0.110	8.05×10^{-2}	1.59×10^{-2}	0.198	T	C	H	C	T	H
55.48	"	"	2.96	2.88	2.42	8.01×10^{-2}	1.65×10^{-2}	0.206	6.30×10^{-2}	2.00×10^{-2}	0.317	4.17×10^{-2}	1.19×10^{-2}	0.285	H	T	C	T	C	H

Unit: particles/cc

Av. 0.167

Av. 0.250

Av. 0.191

Table 3. Correlation of Initial Collection Efficiency

Investigators	Expressions	K_1
Paretsky (1972) [quoted as Doganoglu (1975)]	$(\eta_o)_1 = 2.0 N_{St}^{1.13}$	1.5/ ϵ
Meisen and Mathur (1974)	$(\eta_o)_1 = 0.00075 + 2.6 N_{St}$	
Knettig and Beeckmans (1974)	$(\eta_o)_1 = 3.76 \times 10^{-3} - 0.454 N_{St} + 9.68 N_{St}^2 - 16.2 N_{St}^3$ for $N_{St} > 0.0416$	
Maltcher <u>et al.</u> (1978)	$(\eta_o)_1 = \beta N_{St}$	1.5
	$1.0 N_{Re,s} < 100$ $\beta = \{0.01 N_{Re,s} < 100 \quad N_{Re,s} < 1,000$ $10 N_{Re,s} > 1,000$	
Doganoglu (1975)	$(\eta_o)_{1,o} = 2.89 N_{St} + 6.89 N_G$ for $dg = 100 \mu m$ $(\eta_o)_{1,o} = 0.0583 N_{St} N_{Re,s} + 1.42 N_G$ for $dg = 600 \mu m$	1.5
Thambimuthu <u>et al.</u> (1978)	$(\eta_o)_{1,I} = 10^5 N_{St}^3$ for $10^{-3} < N_{St} < 10^{-2}$	1.5
Thambimuthu (1980)	$(\eta_o)_1 = \left(\frac{N_{St}}{N_{St} + 0.062\epsilon} \right)^3$ for $1.26 < N_{Re} < 130$ $2 \times 10^{-3} < N_{St} < 2 \times 10^{-2}$ $0.33 < \epsilon < 0.4$	
Goren (1982)	$(\eta_o)_1 = 1270 N_{St}^{2.25} (N_{St,eff})^{3.55}$	1.5
D'Ottavio and Goren (1983)	$(\eta_o)_1 = \frac{(N_{St,eff})^{3.55}}{1.67 + (N_{St,eff})^{3.55}}$	1.5
	where $N_{St,eff} = [A(\alpha) + 1.14 N_{Re,s}^{1/2} (1-\alpha)^{-3/2}] \frac{N_{St}}{2}$	
	$A(\alpha) = \frac{(6 - 6\alpha^{5/3})}{(6-9\alpha^{1/3} + 9\alpha^{5/3} - 6\alpha^2)}$	
	$\alpha = 1 - \epsilon$	
Gal <u>et al.</u> (1985)	$(\eta_o)_1 = \frac{2(N_{St}')^{3.9}}{4.3 \times 10^{-6} + (N_{St}')^{3.9}}$	1.5
	where $N_{St}' = N_{St} F$ $F = 1.0 + 1.75 N_{Re,s} / [150(1-\epsilon)]$	
Pendse and Tien (1982)	$(\eta_o)_1 = B [N_{St} + 0.48 (4 - 4 \frac{N_R}{d_c^*} - \frac{N_R^2}{d_c^{*2}})^{1/2} \frac{N_R^{1.041}}{d_c^*}]$	
	where $B = 1 + 0.04 N_{Re,s}$ $N_R = dp/dg$	
Yoshida and Tien (1986)	$\eta_o = B [100 N_{St}^2 + 0.19 (4 - \frac{4N_R}{d_c^*} - \frac{N_R^2}{d_c^{*2}})^{1/2} \frac{N_R^{1.041}}{d_c^*}]$	
	for $N_R < 0.002$ $N_{St} < 0.01$	
	$\eta_o = B [N_{St} + 0.48 (4 - \frac{4N_R}{d_c^*} - \frac{N_R^2}{d_c^{*2}})^{1/2} \frac{N_R^{1.041}}{d_c^*}]$	
	for $N_R > 0.002$ $N_{St} < 0.01$	
	$B = 7 - 6 \exp [-0.0065 N_{Re,s}]$	

Table 4. Summary of Comparisons Between Experimental Results with Prediction Based on Different Correlations

Experimental Condition				Experimental Value of η_o		Comparison with correlation, $(\eta_o)_{pred}/(\eta_o)_{exp}$					
N_{St}	N_{Re}	ϵ	dp/dg	$K_1 = \left[\frac{6}{(1-\epsilon)^{2/3}} \right]^{1/3}$	1.5	Thambimuthu	D'Ottavio & Goren	Gal <u>et.</u> al	Pendse & Tien	Yoshida & Tien	
1.05×10^{-3}	1.969	0.327	2.15×10^{-3}	3.30×10^{-3}	3.55×10^{-3}	0.03	0.0015	0.00037	1.84	1.84	
2.11×10^{-3}	3.938	0.327	2.15×10^{-3}	2.82×10^{-3}	3.03×10^{-3}	0.28	0.024	0.0073	2.74	2.74	
2.19×10^{-3}	9.459	0.335	4.47×10^{-3}	5.00×10^{-3}	5.42×10^{-3}	0.16	0.0093	0.0039	2.48	2.48	
3.16×10^{-3}	1.868	0.320	3.98×10^{-3}	8.54×10^{-3}	9.14×10^{-3}	0.28	0.033	0.01	1.49	1.49	
4.39×10^{-3}	1.892	0.335	4.47×10^{-3}	6.55×10^{-3}	7.11×10^{-3}	0.75	0.094	0.048	2.32	2.32	
5.20×10^{-3}	0.399	0.345	1.06×10^{-2}	1.71×10^{-2}	1.87×10^{-2}	0.40	0.04	0.03	1.72	1.72	
5.27×10^{-3}	9.843	0.327	2.15×10^{-3}	3.11×10^{-3}	3.35×10^{-3}	2.64	0.78	0.33	4.41	4.41	
6.32×10^{-3}	3.737	0.320	3.98×10^{-3}	7.84×10^{-3}	8.38×10^{-3}	1.68	0.50	0.19	2.19	2.19	
6.57×10^{-3}	0.897	0.338	8.28×10^{-3}	1.52×10^{-2}	1.65×10^{-2}	0.82	0.14	0.09	1.70	1.70	
1.04×10^{-2}	7.986	0.345	1.06×10^{-2}	2.25×10^{-2}	2.47×10^{-2}	1.43	0.39	0.37	1.56	1.56	
1.05×10^{-2}	19.69	0.327	2.15×10^{-3}	1.70×10^{-2}	1.83×10^{-2}	2.20	2.23	1.54	1.59		
1.10×10^{-2}	4.728	0.335	4.47×10^{-3}	1.68×10^{-2}	1.82×10^{-2}	2.27	1.18	0.79	1.48		
1.32×10^{-2}	1.795	0.338	8.28×10^{-3}	2.77×10^{-2}	3.02×10^{-2}	1.90	0.98	0.79	1.22		
1.56×10^{-2}	0.379	0.344	1.96×10^{-2}	7.58×10^{-2}	8.31×10^{-2}	0.90	0.43	0.50	0.80		
1.58×10^{-2}	9.338	0.320	3.98×10^{-3}	3.07×10^{-2}	3.32×10^{-2}	2.62	3.85	2.26	1.09		
1.76×10^{-2}	1.969	0.328	9.25×10^{-3}	3.02×10^{-2}	3.26×10^{-2}	3.06	3.13	2.27	1.37		
2.20×10^{-2}	9.455	0.335	4.47×10^{-3}	9.20×10^{-2}	9.98×10^{-2}	1.36	2.49	2.53	0.48		
2.60×10^{-2}	19.957	0.345	1.06×10^{-2}	1.02×10^{-1}	1.12×10^{-1}	1.48	2.05	2.68	0.53		
2.76×10^{-2}	1.191	0.334	1.68×10^{-2}	1.39×10^{-1}	1.50×10^{-1}	1.19	1.77	2.02	0.48		
3.11×10^{-2}	0.758	0.344	1.96×10^{-2}	1.42×10^{-1}	1.55×10^{-1}	1.35	2.10	3.17	0.55		
3.16×10^{-2}	18.676	0.320	3.98×10^{-3}	7.88×10^{-2}	8.43×10^{-2}	2.75	8.33	11.65	0.89		
3.29×10^{-2}	4.485	0.338	8.28×10^{-3}	1.68×10^{-1}	1.78×10^{-1}	1.28	2.80	3.81	0.37		
3.53×10^{-2}	3.938	0.328	9.25×10^{-3}	1.20×10^{-1}	1.30×10^{-1}	1.97	4.71	6.07	0.54		
5.20×10^{-2}	3.991	0.345	1.06×10^{-2}	9.24×10^{-2}	1.01×10^{-1}	3.53	8.00		0.95		
5.51×10^{-2}	2.382	0.334	1.68×10^{-2}	2.00×10^{-1}	2.17×10^{-1}	1.72	3.82		0.49		
6.57×10^{-2}	8.970	0.338	8.28×10^{-3}	1.20×10^{-1}	1.30×10^{-1}						
7.78×10^{-2}	1.893	0.344	1.96×10^{-2}	8.39×10^{-2}	9.20×10^{-2}		Bounce-off occurs				
8.70×10^{-2}	0.399	0.352	4.56×10^{-2}	3.06×10^{-1}	3.38×10^{-1}						
8.82×10^{-2}	9.843	0.328	9.25×10^{-3}	9.56×10^{-2}	1.03×10^{-1}						
1.38×10^{-1}	5.952	0.334	1.68×10^{-2}	7.09×10^{-2}	7.69×10^{-2}						
1.56×10^{-1}	3.786	0.344	1.96×10^{-2}	6.16×10^{-2}	6.75×10^{-2}						
1.74×10^{-1}	0.799	0.352	4.56×10^{-2}	2.42×10^{-1}	2.68×10^{-1}						
1.76×10^{-1}	19.686	0.328	9.25×10^{-3}	3.02×10^{-2}	3.26×10^{-2}						
2.76×10^{-1}	11.90	0.334	1.68×10^{-2}	2.61×10^{-2}	2.83×10^{-2}						
4.35×10^{-1}	1.996	0.352	4.56×10^{-2}	1.13×10^{-1}	1.25×10^{-1}						
8.70×10^{-1}	3.991	0.352	4.56×10^{-2}	8.60×10^{-2}	9.49×10^{-2}						

Table 5

Comparison of Experimental Conditions of the Present Study
and Those Used by D'Ottavia and Goren

Our Experiment		Goren's Experiments	
grain	glass beads	glass beads	alumina beads
grain size (dg)	103 μm , 244 μm , 508 μm	4 mm or 4,000 μm	2mm or 2,000 μm
Aerosol Particles	Uniform Latex Particles (polyvinyl toluene)	KHP (solid aerosol) DOP (liquid aerosol)	KHP (solid aerosol) DOP (liquid aerosol)
Aerosol particle size (dp)	1.0, 1.091, 2.02 μm 4.1, 4.7 μm	for KHP 0.59, 1.14, 2.27 μm	for KHP 0.68, 1.17, 2.28 μm 4.54 μm
$dp/dg = N_R$	$2.15 \times 10^{-3} \sim 4.56 \times 10^{-2}$	$1.48 \times 10^{-4} \sim 5.68 \times 10^{-4}$	$3.4 \times 10^{-4} \sim 2.27 \times 10^{-3}$
Filter bed Diameter	1.5" ID Plexiglass	2" ID Copper Tube	2" ID Copper Tube
Bed Height (L)	0.14 $\sim$ 0.69 cm	3 cm $\sim$ 10.1 cm	2.8 cm $\sim$ 10.1 cm
Superficial gas Velocity (V_s)	5.55 $\sim$ 55.48 cm/sec	13 $\sim$ 620 cm/sec	10 $\sim$ 380 cm/sec
N_{Re}	0.399 $\sim$ 19.96	32 $\sim$ 1620	13 $\sim$ 500
N_{St}	$1.05 \times 10^{-3} \sim 8.70 \times 10^{-1}$	$4.3 \times 10^{-4} \sim 2.05 \times 10^{-2}$	$8.5 \times 10^{-5} \sim 8.5 \times 10^{-2}$
$N_{St_{eff}}$	0.036 $\sim$ 27.5	0.03 $\sim$ 12	0.017 $\sim$ 9.5
Instrument for Concentration Measurement	TSI, H/R, Climet	Climet	Climet

FIG. 1 SCHEMATIC DIAGRAM OF EXPERIMENTAL APPARATUS

FIG. 2a INFLUENT OR EFFLUENT CONCENTRATION vs. TIME

FIG. 2b CONCENTRATION RATIO vs. TIME

FIG. 3a INFLUENT OR EFFLUENT CONCENTRATION vs. TIME

FIG. 3b CONCENTRATION RATIO vs. TIME

FIG. 4 VALUES OF THE INITIAL COLLECTION EFFICIENCY OBTAINED FOR A GIVEN EXPERIMENTAL RUN

FIG. 5 VALUES OF THE INITIAL COLLECTION EFFICIENCY OBTAINED BY A GIVEN EXPERIMENTAL RUN

FIG. 6a $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING CLIMET COUNTER

FIG. 6b $\sigma_{\bar{\eta}_0}/\bar{\eta}_0$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING HIAC/ROYCO COUNTER

FIG. 6c $\sigma_{\bar{\eta}_O}/\sigma_{\bar{\eta}_O}$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING TSI PARTICLE SIZER

FIG. 7a $\sigma_{\bar{C}_{in}} / \bar{C}_{in}$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING CLIMET COUNTER

FIG. 7b $\sigma_{\bar{C}_{in}}/\bar{C}_{in}$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING HIAC/ROYCO COUNTER

FIG. 7c $\sigma_{\bar{C}_{in}}/\bar{C}_{in}$ vs. $\bar{C}_{in}$ FOR DATA OBTAINED USING TSI PARTICLE SIZER

FIG. 8 SINGLE COLLECTOR EFFICIENCY vs. $N_{st_{eff}}$

FIG. 9 INCORPORATION OF N_R INTO THE CORRELATION OF η_0

FIG. 10 ADHESION PROBABILITY AS A FUNCTION OF $N_{st\,eff}$

Part II.

Deposition of Polydispersed Aerosol Particles in Granular Media

Introduction

For most practical applications of granular filtration, the gas streams to be treated inevitably contain particulates of various sizes. Thus it is of considerable practical importance to develop predictive capability for predicting the changes in collection efficiencies of these particles as the extent of deposition increases.

This particular phase of investigation is concerned with collecting experimental data on the filtration of polydispersed suspensions. The particle concentrations of particles of all sizes in both the influent and effluent of the experimental filter were determined as a function of time. This study may be considered as a generalization of the earlier work of Takahashi et al. which was concerned with the filtration of monodispersed suspensions.

Experimental Work

The experimental apparatus, systems and procedures are identical to those used in the earlier studies (Walata et al., 1986; Takahashi et al., 1986) and therefore will not be described. The only modifications of the apparatus were the following: (a) the test suspensions were prepared by mixing latex particles of three different sizes ($d_p = 1.091 \mu\text{m}$, $2.02 \mu\text{m}$ and $4.7 \mu\text{m}$) so that the test aerosols were polydispersed; and (b) the particle concentrations were determined using a HIAC/Royco model PA-720 particle counter.

A large number of experiments were conducted using three different grain sizes ($d_g = 508$, 244 and $103 \mu\text{m}$) and gas velocity up to 60 cm/s . For each experiment, influent and effluent concentrations and pressure drop were obtained at various times.

Results

The experimental data were analyzed in the same manner as used by Walata et al. (1986), namely, the effect of deposition on the collection efficiency and pressure gradient can be expressed as

$$F_1 = \frac{\eta}{\eta_o} = 1 + \alpha_1 \sigma^{\alpha_2} \quad (1)$$

$$F_2 = \frac{(\partial p / \partial z)}{(\partial p / \partial z)_o} = 1 + \beta_1 \sigma^{\beta_2} \quad (2)$$

where α_1 , α_2 , β_1 and β_2 are empirical constants. Walata suggested an extrapolation procedure for obtaining the relationships of η/η_o vs. σ and $\frac{\partial p / \partial z}{(\partial p / \partial z)_o}$ vs. σ from concentration and pressure drop measurements. His results also indicate that one may use the uniform deposition assumption to obtain F_1 (or F_2) vs. σ relationship if $L/d_g < 20$ where L and d_g are the filter height and grain diameter, respectively. This approximate procedure was used in the present work.

Once the F_1 (or F_2) vs. σ results were obtained, attempts were made to see whether these relationships can be described by Equation (1) [or Equation (2)]. These attempts, however, were found unsuccessful. The F_1 vs. σ was found to exhibit rather complex behavior as shown in Fig. 1. In Fig. 1, the value of η for $d_p = 1.091, 2.02$ and $4.7 \mu m$ vs. $\bar{\sigma}$ calculated from the influent and effluent concentration data obtained at two different superficial velocities are shown. For the smallest particles, ($d_p = 1.091 \mu m$), the η vs. σ relationship displayed a certain constancy. For particles with diameter

$d_p = 2.02 \mu\text{m}$, the superficial gas velocity was found to play an important role in determining the dependence of η on σ . On the other hand, the collection efficiency of the largest particle ($d_p = 4.7 \mu\text{m}$) was found to reach unity very quickly, indicating that the retention of particles of this size, with $\sigma = 2 \times 10^{-4}$, was primarily dense to straining.

A systematic comparison of the η vs. σ data of polydispersed suspensions with the η vs. σ relationship established from monodispersed suspension data (Takahashi et al., 1986) was made and a summary of the comparison is given in Table 1. It is clear that the results previously obtained cannot be extended to polydispersed suspensions. Without an adequate theoretical understanding, a proper interpretation of the polydispersed suspension data would be difficult to achieve. This fact deserves special attention in future studies dealing with deposition of polydispersed aerosols.

Table 1

Summary of the Results of the Increase in Collection Efficiency, η ,
Due to Deposition of Polydispersed Suspensions With Monodispersed Suspensions

$d_g = 103 \mu\text{m}$

$u_s = 14.62 \text{ cm/s}$

$u_s = 29.24 \text{ cm/s}$

$u_s = 58.48 \text{ cm/s}$

$d_p = 1.091 \mu\text{m}$

similar

less increase

much less increase

$d_p = 2.02 \mu\text{m}$

less increase

higher increase

-

$d_p = 4.7 \mu\text{m}$

unknown

unknown

unknown

$d_g = 244 \mu\text{m}$

$d_p = 1.091 \mu\text{m}$

less increase

similar

similar

$d_p = 2.02 \mu\text{m}$

similar

similar

higher increase

$d_p = 4.7 \mu\text{m}$

unknown

unknown

unknown

$d_g = 508 \mu\text{m}$

$d_p = 1.091 \mu\text{m}$

similar

similar

-

$d_p = 2.02 \mu\text{m}$

higher increase

higher increase

-

FIG. 1 COLLECTION EFFICIENCIES VS $\bar{\sigma}$ FOR THREE PARTICLE SIZES
(Data collected using polydispersed suspension)

Notations

d_g	grain diameter
F_1, F_2	defined by equations (1) and (2), respectively
L	filter height
p	pressure
z	axial distance

Greek Letters

α_1, α_2	constants of Equation (1)
β_1, β_2	constants of Equation (2)
η	unit collector efficiency
η_o	initial value of η
σ	specific deposit

References

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Part III.

In Situ Observation of Aerosol Deposition
in a Two-Dimensional Model Filter

Introduction

The use of model filters in aerosol research was pioneered by Fuchs and Stechkina (1963). In contrast to ordinary filters, the collecting elements of a model filter are often of uniform size and arranged with simple and regular patterns. As a result, theoretical analyses of aerosol filtration in model filters can be conducted rigorously. One can therefore make comparisons between theory and experiments of model filters in a more decisive manner than ordinary filters.

Another potential advantage of model filters is that they can be often constructed in a way permitting direct and detailed observations of the filtration process, especially the manner and the extent of particle deposition. This potential advantage, however, has hardly been exploited by aerosol workers even though model filters have been used in hydrosol filtration by a number of investigators for this purpose. Beginning with the work of Stein (1940), several studies (Moraudas and Eisenklam, 1965, Ison and Ives, 1964, Payatakes et al., 1981) have employed the use of specially-built filters to examine the morphology of particle deposit formed in liquid filtration. The work of Payatakes et al. is particularly significant on two accounts. The dimension of the collecting elements in their model filter is comparable to real filter grains; and the method used for observation is non-intrusive, continuous and capable of covering the entire filter media domain. Detailed information about the mechanics of particle deposition and reentrainment of hydrosol filtration with or without polyelectrolytes were obtained by Payatakes et al. (1981).

In a recent study, Ushiki and Tien (1984) demonstrated that a two-dimensional filter of a design similar to that of Payatakes et al. can be applied in aerosol filtration studies as well. The present work represents a further extension of

the Ushiki and Tien study in exploring the use of the model filter for aerosol research. Two specific features of this work are: (a) the collecting element of the model filters are arranged randomly rather than regularly as before, in order to obtain a more realistic representation of granular filters, and (b) in addition to qualitative observations, quantitative data on the increase in pressure drop during filtration as a function of the extent of deposition were obtained and compared with those of ordinary granular filters.

Design of Model Filters

The model filter used in the study is composed of a photoetched glass plate and a cover plate. The collecting elements formed on the photoetched plate are rows of large numbers of small cylindrical rods of 688 μm in diameter and 44 μm in height. An enlarged pattern indicating the placement of the collecting elements was first prepared. The enlarged pattern was then photoreduced to the actual dimension. The procedure used in generating the random pattern is described as follows.

Consider a plate with a dimension of $\ell_1 \times \ell_2$ and with a specified bed porosity of ϵ , the total number of cylindrical rods present, n , is

$$n = \frac{4 \ell_1 \ell_2 (1 - \epsilon)}{\pi d_c^2} \quad (1)$$

where d_c is the diameter of the rod.

The generation of the required pattern is to place the n circles on the rectangle of $\ell_1 \times \ell_2$. First the points (center points) within a rectangle of $\ell_1 \times \ell_2$, where the n circles should be placed in order to create a centered hexagonal pattern are identified. A circle of diameter Δd , where Δd is the difference between the distance between any two adjacent center points and the diameter of

the collector, is made around each of these points. Each of these circles is a domain of placement for placing a collector. The random pattern is created by placing one circle of diameter d_c with its center chosen randomly, within each of these domains of placement. For this purpose, a sequence of n pairs of random numbers (a_i, b_i) , $i = 1, 2, \dots, n$ with $0 \leq a_i \leq 1$ and $0 \leq b_i \leq 1$ can be generated. The position of the i -th collector within its domain of placement, expressed by the polar coordinates (r_i, θ_i) is given as

$$r_i = (a_i) \frac{\Delta d}{2} \quad (2)$$

$$\theta_i = 2\pi b_i \quad (3)$$

A major question which needs to be answered is how adequately does the two dimensional filter portray real filters? To provide some idea, the following information are presented. First, for a given row of collectors in the two dimensional filter, the narrowest distance between two adjacent collectors, based on the constricted tubed model of porous media, can be considered as equivalent to a pore constriction in a real filter. Thus from a given design pattern a distribution function for the pore constrictions can be determined as a function of the constriction dimension. Let $F(\Delta d)$ be the number fraction of constrictions whose dimension is equal to or less than Δd . Fig. 1 and 2 shows the relationship of $F(\Delta d)$ vs. Δd for model filters corresponding to two different sets of values d_c and ε . Also included in these two figures are experimental data obtained previously by Payatakes (1973) for beds of nearly monosized glass beads. In Fig. 1, there is no difference between the data of Payatakes and the results of the two dimensional filter. In Fig. 2, in spite of the difference between the two sets of results, the qualitative similarity between them is nevertheless remarkable.

The random pattern used to place the collectors in the model filter showed that there is practically no contact between collectors of a given row. This lack of contact among collectors in a two-dimensional filter is often considered as one of its major deficiencies in a realistic representation of granular media. The collectors of a given row in a two-dimensional filter is equivalent to the filter grains of a given bed height. A micrograph of a cross section of a sand-media ($d_g = 600 \mu\text{m}$) is shown in Fig. 3 (Nye, 1985). It becomes obvious that contact between grains at the same bed height are indeed rare. The major difference between a real granular medium and a two-dimensional model filter as shown in Figs. 1, 2 and 3 is that the pore size distribution of the former covers a wider range than that of the latter.

Experimental Work

Apparatus. A schematic diagram of the apparatus used in this work is shown in Fig. 4. The apparatus is essentially the same as that used by Ushiki and Tien (1984). Briefly speaking, the model filter consists of a photoetched glass plate (filter plate) and a cover plate. The cover plate, in turn, was composed of two separate parts: a smaller upper part for the filter inlet and a larger lower part for the filter and the exit channel. The reasons that the cover plate was made of two separate parts were two-fold. First with this arrangement, it was possible to deepen the filter inlet ($94 \mu\text{m}$ as compared with $44 \mu\text{m}$ in the filter) and, therefore, minimize the possibility of inlet clogging. Second, with the use of two separate parts, a more complete contact between the cover and filter plate could be made and therefore eliminates gas leakage. These plates were put together using fasteners and mechanical springs. Pressure taps were placed at certain depths into the model (corresponding to the fifth, tenth, fifteenth, twentieth and twenty-fifth rows of the collector). The dimensions of the model filter are shown in Fig. 5.

The test aerosols were prepared by dispersing aluminum oxide (Buehler Micropolish No. 40-6610-004, 3.0 μm nominal size) in a nitrogen gas stream with the use of a fluidized bed generator. Cylinder nitrogen gas was passed through the generator into which aluminum oxide powders were placed. The fluidized bed generator had a tapered section to minimize the bed height change during experiments. The test aerosols passed through a mechanical disperser and a settling chamber before entering into the model filter. The purpose of the disperser was to de-agglomerate aluminum oxide particle clusters present in the test aerosol. Oversize particles tended to settle in the settling chamber. With this arrangement, the test aerosols became less polydispersed.

The observation-recording of deposition taking place within the model filters was carried out by using a video-camera (Hitachi HV-620) a video recorder (Sony AV-3600) and television camera (Hitachi VM-A.O.U). The video camera was attached to an optical microscope (Bausch and Lomb) with four lenses of different magnifications.

Procedures. Before each experiment, the components of the model filters were washed in a sonic dismembrator (Artek Systems Co. Model 150) with distilled water. Afterwards, the model filter was assembled and placed on the platform (see Fig. 5). Aluminum oxide powders were placed in the fluidized-bed generator and test aerosols were prepared. The test suspension was then passed through the model filter at specified flow rates.

The video camera, monitor and video recorder were connected to the microscope and the microscope was focused. Continuous observation and recording were made of the deposition process taking place at certain locations of the model filter. The experiments lasted, on the average, approximately 4 or 5 hours.

Conditions. The experiments were conducted using one type of aerosol (aluminum oxide), the size distribution of which is shown in Fig. 6. Three different gas

velocities and several different types of model filters (with $d_c = 688 \mu\text{m}$), namely model filters with collectors arranged in centered hexagonal, square in addition to the random pattern mentioned earlier, were used. The specific conditions of each experiment are listed in Table 1.

Results

Pattern of Deposition. In Figs. 7, 8, 9, 10 and 11, the extent of particle deposition corresponding to the five cases listed in Table 1 are shown. The major conclusions are

(a) Most of the deposition of the model filters took place within the first three or four rows of collectors.

(b) For a given collector, deposition was confined primarily in the neighborhood near the front stagnation point of the collector.

(c) As a natural consequence of (b), the amount of deposition in the model filter with collectors arranged in a square pattern is slight as compared with the other two patterns.

(d) For model filters with collectors arranged in a centered hexagonal pattern, the extent of deposition of collectors of a given row exhibits remarkable similarities in both the manner with which particle deposits are formed and the ultimate shape of the deposit.

(e) In contrast to the behavior of (d), for model filters with random patterns, there is no similarity in deposition patterns among collectors in a given row. Collectors with significant deposition are those situated with comparatively large pore constrictions above them. The gradual pattern of deposition (from row to row moving downward) observed by Ushiki and Tien (1984) was not observed in the random pattern model filters.

The observations that significant deposition occurs in the front stagnation point region of a collector for collectors situated below relatively large pore

constrictions add credence to the speculation raised earlier (Ushiki and Tien, 1984): namely, on a local basis, aerosol deposition may be better described as a two dimensional jet of aerosols impinging upon a collector. However, the impinging jet flow field has never been considered in the trajectory calculation efficiencies.

Pressure Drop of Clean Model Filters. The pressure drop through the model filters was measured for model filters with random, centered hexagonal and square patterns (all with $\epsilon = 0.325$). The flow rates were so varied to give a Reynolds number (N_{Re_s}) value from 2 to 20. If the model filter media are viewed as porous media, the commonly acceptable way of representing the pressure drop-flow rate relationship with the friction factor - Reynolds number correlation can be applied. The friction factor is defined as

$$f_s = \frac{-\Delta P}{2\rho u^2 L} d_c \quad (4)$$

The results are shown in Fig. 12 in the form of f_s vs. N_{Re_s} for the three different types of model filters. Also included in these figures are the relationship of f_s vs. N_{Re_s} according to a number of well-established correlations such as

Blake-Kozeny's Equation
$$f_s = \frac{1}{N_{Re_s}} \frac{150 (1 - \epsilon^2)}{2\epsilon^3} \quad (5)$$

Ergun's Equation
$$f_s = \frac{1}{N_{Re_s}} \left[\frac{150}{2} \frac{1 - \epsilon^2}{\epsilon^3} \right] + \frac{1.75}{2} \frac{1 - \epsilon}{\epsilon^3} \quad (6)$$

Happel's Equation
$$f_s = \frac{9\gamma^3}{N_{Re_s}} \left[\frac{3 + 2\gamma^5}{3 - \frac{9}{2}\gamma + \frac{9}{2}\gamma^5 - 3\gamma^6} \right], \gamma = (1 - \epsilon)^3 \quad (7)$$

Equation of MacDonald et al. (1979)

$$f_s = \frac{1}{N_{Re_s}} \left[\frac{180}{2} \frac{(1 - \epsilon)^2}{\epsilon^3} \right] + \frac{1.8}{2} \frac{(1 - \epsilon)}{\epsilon^3} \quad (8)$$

as well as the results calculated with the use of the constricted tube model by Payatakes et al. (1973) and by Venkatesan and Rajagopalan (1979) for the parabolic and hyperbolic wall geometry, respectively. The friction factor obtained from the model filters, for a given N_{Re_s} is consistently higher than any one of the above-mentioned expressions. Among the three types of model filters, the results obtained from the random and centered hexagonal patterns are sufficiently close. The square pattern model filter gave the lowest value of f_s among the three model filters used, but yielded results similar to those predicted by either Payatakes et al. or Venkatesan and Rajagopalan. Since the flow channel formed in a model filter with square pattern is rather similar to that of a periodically constricted tube, the results are therefore not surprising.

Pressure Drop Increase Due to Deposition. One of the characteristic features of granular filtration is the increase in pressure drop necessary to maintain a given flow rate through a filter as the filter becomes clogged. The same behavior was also displayed by the model filters as shown in Figure 13. In Fig. 13, the values of $\Delta P/\Delta P_0$, where the pressure difference, ΔP , was measured in reference to point A as shown in Fig. 5 and ΔP_0 is the value of ΔP when the filter clean, are plotted against the depth of the filter at various times for a model filter with collectors arranged in random patterns. Similar results were also obtained with model filters with collectors arranged in centered hexagonal pattern.

Since the two dimensional model filters used in this study allow a direct and continuous observation of the extent of deposition, it becomes possible to relate the pressure drop increase with the extent of deposition. As shown in Fig. 13, most of the pressure drop occurred across the first five rows of collectors. The extent of the deposition can be expressed by the apparent specific deposit, $\hat{\sigma}$, defined as the volume of deposits per unit volume of bed. Note that $\hat{\sigma}$ is not the same as σ , the specific deposit defined as the amount of particle (in volume) per unit volume of bed. $\hat{\sigma}$ and σ are related by the expression

$$\hat{\sigma} = \sigma / \epsilon_d \quad (9)$$

where ϵ_d is the porosity of the deposit.

The micrographs of the collectors taken during the course of filtration identify the space occupied by deposited aerosols as darkened area around the collector. Accordingly, $\hat{\sigma}$ can be estimated to be

$$\hat{\sigma} = \frac{\text{Area occupied by deposited aerosols}}{\text{Total area of the collectors}} \quad (10)$$

From prints of the micrographs, one may cut off the area occupied by the deposited aerosols around the collector of the first five rows and obtain its weight. The ratio of this weight to the weight corresponding to the entire space occupied by the first five rows of collectors given the value of $\hat{\sigma}$. In Fig. 14, the dimensionless pressure drop increase across the first four rows are plotted against $\hat{\sigma}$ for results obtained using model filters with collectors in two different arrangements (centered hexagonal and random) two porosities (0.325 and 0.45) and at two different velocities ($u = 41.7$ cm/s and 59.6 cm/s). In spite of significant differences in experimental conditions, the data points displayed remarkable consistency. An empirical correlation of the following form was obtained

$$\frac{(\Delta P)_{AB} - (\Delta P)_0}{\Delta P_0} = 24.73 \hat{\sigma}^{1.247} \quad (11)$$

where $(\Delta P)_{AB}$ is the pressure over the two locations as shown in Fig. 5.

As shown previously (Tien and Payatakes, 1979; Tien, 1981), the effect of particle deposition on pressure drop can be expressed in terms of the changes in pressure gradient as functions of

$$\frac{(\partial P / \partial Z)}{(\partial P / \partial Z)_0} = F_2(\sigma) \quad (12)$$

As an approximation, the left hand side of Equation (11) can be considered to be $\frac{(\partial P / \partial Z)}{(\partial P / \partial Z)_0} - 1$. With the approximation together with the use of Equation (8),

Equation (10) may be transformed into

$$\frac{(\partial P/\partial Z)}{(\partial P/\partial Z)_0} = 1 + 24.73 \hat{\sigma}^{1.247} \quad (13)$$

More recently, Walata (1984) examined the transient behavior of granular aerosol filtration experimentally. The correlation of his pressure drop data obtained from filter bed composed of filter grains with 505 μm in diameter and using 2.02 μm latex particles aerosols at a superficial velocity of 22.6 cm/s is found to be

$$\frac{(\partial P/\partial Z)}{(\partial P/\partial Z)_0} = 1 + 630 \hat{\sigma}^{0.945} \quad (14)$$

An exact comparison between the above two expressions, of course, is not possible, since the value of ϵ_d is unknown. It is obvious that Equation (13) predicts a smaller increase in $(\partial P/\partial Z)/(\partial P/\partial Z)_0$ than Equation (14). This underestimation is not expected since Equation (13) is based on approximating $(\partial P/\partial Z)/(\partial P/\partial Z)_0$ by the total pressure drop across five rows of collectors.

Another comparison of Equation (13) with previous work is shown in Fig. 15. Pendse and Tien (1982) applied the stochastic modelling to granular filtration and considered that the increase in pressure drop is caused by the drag forces acting on the deposited particles. The results according to Equation (13) are also shown in Fig. 5 assuming that ϵ_d is 0.65. The behavior of the model filter was found to be consistent with the results of stochastic simulations.

Table 1. Experimental Condition

Porosity $\epsilon(-)$	Arrangement	Particle Diameter $D_p (\mu\text{m})$	Superficial Velocity $V_s (\text{cm/s})$	Case
0.325	Hexagonal	3	22.37	A
0.325	Random R	3	41.7	B
0.325	Random R	3	59.6	C
0.325	Normal N	3	59.6	D
0.45	Random R	3	59.6	E

FIG. 1 Constriction Diameter Distribution

FIG. 2 Constriction Diameter Distribution

FIG. 3

MICROGRAPH (4.5x) OF A THIN SECTION OF SAND BED ($d_g = 600 \mu m$)
CONTACTS BETWEEN SAND GRANULES ARE PRACTICALLY ABSENT

FIG. 4 SCHEMATIC DIAGRAM OF EXPERIMENTAL APPARATUS

FIG. 5 Test Section of Model Filter

FIG.6 SIZE DISTRIBUTION OF TEST AEROSOL PARTICLES

$t = 40 \text{ min}$

$t = 155 \text{ min}$

$t = 255 \text{ min}$

$t = 343 \text{ min}$

$t = 396 \text{ min}$

FIG. 7 DEPOSITION PATTERN OF CASE (A). TABLE 1

$t=45 \text{ min}$

$t=95 \text{ min}$

$t=174 \text{ min}$

$t=255 \text{ min}$

$t=318 \text{ min}$

FIG. 8 DEPOSITION PATTERN OF CASE (B), TABLE 1

$t=28$ min

$t=54$ min

$t=132$ min

$t=147$ min

$t=225$ min

$t=251$ min

$t=276$ min

FIG. 9 DEPOSITION PATTERN OF CASE (C) TABLE 1

$t = 23 \text{ min}$

$t = 43 \text{ min}$

$t = 79 \text{ min}$

$t = 143 \text{ min}$

$t = 192 \text{ min}$

$t = 254 \text{ min}$

$t = 304 \text{ min}$

FIG. 10 PATTERN OF DEPOSITION CASE(D) TABLE 1

$t = 49 \text{ min}$

$t = 165 \text{ min}$

$t = 266 \text{ min}$

$t = 347 \text{ min}$

$t = 437 \text{ min}$

$t = 474 \text{ min}$

FIG. 11 DEPOSITION PATTERN OF CASE(E) TABLE 1

FIG.12 Relation between Friction Factor and Flow Reynolds Number

FIG. 13 Relation between Pressure Drop Ratio and Filter Thickness(case P)

FIG.14 Experimental Results of Pressure Drop for Model Filter

FIG. 15 Experimental Results of Pressure Drop compared with Liquid Filtration Data

Nomenclature

a_i, b_i	random numbers between 0 and 1
d_c	collector diameter
f_s	friction factor
$F(\Delta d)$	pore constriction distribution function
F_2	defined as $(\partial P / \partial Z) / (\partial P / \partial Z)_0$
L	filter height
ℓ_1, ℓ_2	dimensions of model filter
n	number of collectors in a model filter
r_i	radial coordinate
u	superficial velocity

Greek Letters

Δd	diameter of the domain of placement
ΔP	pressure drop
ϵ	porosity
ϵ_d	deposit porosity
ρ	density of fluid
$\bar{\sigma}$	specific deposit
$\hat{\sigma}$	apparent specific deposit
μ	viscosity of fluid

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Part IV.

Solid Velocity in Cross-Flow Granular Moving Bed

Introduction

Cross-flow moving-bed technology has been applied in a number of gas-solid contacting operations involving chemical reaction, mass transfer, and separation. Two major examples of its use in the chemical and process industries are cross-flow granular aerosol filtration and regeneration of reforming catalysts.

The major advantage of conducting gas-solid contacting in moving beds is that the operations can be continuous. While fixed-bed operations are inherently unsteady-state, with operations inevitably ending after a certain time, moving beds may be operated co-currently, counter-currently, or in a cross-flow configuration. Cross-flow moving beds also have the advantage that the pressure drop for maintaining gas flow is rather modest even for cases of large amounts of solids.

The cross-flow configuration does, however, suffer from a major defect: dissymmetry. Because of the gas drag exerted on the solids, the compressive stress in the solid phase increases along the direction of the gas flow. Consequently, the flat velocity profile approximation used for solid flow in counter-current and co-current moving beds does not always apply. In the extreme, if the pressure gradient along the gas-flow direction is sufficiently large, solid particles may become pinned against the downstream wall of the bed (see Fig. 1 for illustration), thus disrupting or even totally stopping the solid movement.

Study of the behavior of the cross-flow moving bed began only recently. Bridgwater (1981) commented in a general way on the problem of pinning.

Ginestra and Jackson (1985) and Doyle, Jackson and Ginestra (1986) examined, in detail, the solid pinning that occurred in a rectangular moving bed and in an angular moving bed, respectively. In a recent work, Mizukami et al. (1984, 1985) measured and reported on solid velocity in a model cross-flow moving-bed filter. As could be expected, solid velocity is an important variable in determining the performance of cross-flow moving-bed filters. It strongly affects the way particle deposits are formed in the filter media, which, in turn, strongly affects the pressure drop increase attributable to clogging of the filter media.

In the present study, a simple analysis was conducted of solid movement in cross-flow moving-bed filters. The analysis was done by obtaining the stress distribution throughout the solid phase and by comparing these stresses with the frictional stress between particles and between the particle and the surface of the filters. Based on the results of this analysis, the solid velocity profile can be predicted if one knows the imposed gas pressure gradient and the material constant of the solid substance.

Stress Distribution

The velocity profile of the granular particles in a cross-flow moving bed can be estimated by comparing the shear stress in the solid phase with both the internal frictional stress between the granules and the wall frictional stresses. Accordingly, the first step of the analysis is to obtain the stress distribution in the solid phase, a process based on the fact that the granular particles can be treated as a continuum. Furthermore, it is assumed that

- (a) the void fraction of the moving bed is a constant;

(b) the relevant forces are gas drag, interparticle force, gravity, and wall-particle force; and

(c) the gas flow is one-dimensional.

The two-dimensional stress equations are

$$\frac{\partial \sigma_{xx}}{\partial x} + \frac{\partial \tau_{yx}}{\partial y} = - (\partial p / \partial x) \quad (1)$$

$$\frac{\partial \sigma_{yy}}{\partial y} + \frac{\partial \tau_{xy}}{\partial x} = \pm \gamma \quad (2)$$

where σ_{xx} and σ_{yy} are the compressive stress components and τ_{yx} and τ_{xy} are the shear stress components. τ_{yx} and τ_{xy} are considered to be the same. $(\partial p / \partial x)$ is the pressure gradient along the x-direction, which maintains the gas flow, and γ , the weight of the granular substance per unit volume. Equations (1) and (2) are written with reference to Fig. 1. The sign of Equation (2) depends upon the location of the x-coordinate. Near $x = 0$, the negative sign applies; near $x = L$, the positive sign applies.

By using L , the thickness of the filter medium, and γL as the normalizing factors for the spatial variables (x and y) and the stresses, respectively, one has

$$\frac{\partial \sigma_{xx}^*}{\partial x^*} + \frac{\partial \tau_{xy}^*}{\partial y^*} = - \pi^* \quad (3)$$

$$\frac{\partial \sigma_{yy}^*}{\partial y^*} + \frac{\partial \tau_{xy}^*}{\partial x^*} = \pm 1 \quad (4)$$

where

$$x^* = x/L, \quad y^* = y/L$$

$$\sigma^* = \sigma/\gamma L \quad \tau^* = \tau/\gamma L \quad (5)$$

$$\pi^* = (-\partial P/\partial x)/\gamma$$

If the end effect is ignored (i.e., the y-dependence), Equations (3) and (4) are reduced to

$$\frac{\partial \sigma_{xx}^*}{\partial x} = \pi^* \quad (6)$$

$$\frac{\partial \tau_{xy}^*}{\partial x} = \mp 1 \quad (7)$$

The solutions to Equations (6) and (7) depend upon the boundary conditions, namely, the stresses acting at the two walls ($x=0$ and L). Referring to Fig. 1, and letting $\sigma_0 = \sigma_{xx}|_{x=0}$, $\sigma_L = \sigma_{xx}|_{x=L}$, $\tau_0 = \tau_{yy}|_{x=0}$, and $\tau_L = \tau_{xy}|_{x=L}$, one finds the force balance along the x- and y-directions to yield

$$\sigma_L - \sigma_0 = (-dp/dx) L \quad (8)$$

$$\frac{\partial \sigma_{yy}}{\partial y} = \gamma - (\tau_0 + \tau_L)/L \quad (9)$$

The compressive and shear stresses at the wall are related to one another by the following expressions:

$$\text{At } x = 0, \quad \tau_{xy} = \mu_{w_0} \cdot \sigma_{xx} \quad \text{or} \quad \tau_0 = \mu_{w_0} \sigma_0 \quad (10)$$

$$\text{At } x = L, \quad \tau_{xy} = \mu_{w_L} \sigma_{xx} \quad \text{or} \quad \tau_L = \mu_{w_L} \sigma_L \quad (11)$$

where μ_{w_0} and μ_{w_L} are the wall friction coefficient. Furthermore, at $x = 0$, if the stresses are in the critical state, one has

$$\sigma_0 = k \sigma_{yy} \quad (12)$$

Combining Equations (8), (9), (10), (11), and (12) and writing the final equation in the dimensionless form, one has

$$\frac{\partial \sigma_{yy}^*}{\partial y^*} = 1 - \mu_{w_L} \pi^* - k \mu_{w_T} \sigma_y^* \quad (13)$$

where

$$\mu_{w_T} = \mu_{w_0} + \mu_{w_L} \quad (14)$$

The solution to Equation (13) is subject to the boundary condition

$$\sigma_{yy}^* = \sigma_{yy_0}^* \quad \text{at} \quad y^* = 0 \quad (15)$$

which is given as

$$\sigma_{yy}^* = \frac{1 - \mu_{w_L} \pi^*}{k \mu_{w_T}} (1 - e^{-k \mu_{w_T} y^*}) + \sigma_{yy_0}^* e^{-k \mu_{w_T} y^*} \quad (16)$$

For large y^* (namely, by ignoring the end effect), one has

$$\sigma_{yy}^* = \frac{1 - \mu_{w_L} \pi^*}{k \mu_{w_T}} \quad (17)$$

From Equations (17), (12), (8), (10), and (11), the stresses at the two walls are:

$$\text{at } x^* = 0, \quad \sigma_{xx}^* = \sigma_o^* = \frac{1 - \mu_{wL} \pi^*}{\mu_{wT}} \quad (18 \text{ a})$$

$$\tau_{xy}^* = \tau_o^* = \frac{\mu_{wo}}{\mu_{wT}} (1 - \mu_{wL} \pi^*) \quad (18 \text{ b})$$

$$\text{at } x^* = 1 \quad \sigma_{xx}^* = \sigma_L^* = \pi^* + \frac{1 - \mu_{wL} \pi^*}{\mu_{wT}} = \frac{1 + \mu_{wo} \pi^*}{\mu_{wT}} \quad (19 \text{ a})$$

$$\tau_{xy}^* = \tau_L^* = \frac{\mu_{wL}}{\mu_{wT}} (1 + \mu_{wo} \pi^*) \quad (19 \text{ b})$$

The stress distribution (σ_{xx}^* vs. y , τ_{xy}^* vs. y) can be found from the solutions to Equations (6) and (7) subject to the boundary conditions of Equations (18) and (19). However, it should be also noted that the normal stresses (σ_{xx}^* and σ_{yy}^*) cannot be negative. This condition is met if σ_o^* is always positive or zero, or if $\pi^* \leq 1/\mu_{wL}$. Thus, for $\pi^* \leq 1/\mu_{wL}$, σ_{xx}^* and σ_{xy}^* are found to be

$$\sigma_{xx}^* = \sigma_o^* + \pi^* x^* = \frac{1}{\mu_{wT}} + \left(x^* - \frac{\mu_{wL}}{\mu_{wT}}\right) \pi^* \quad (20)$$

$$\tau_{xy}^* = \tau_o^* - x^* \quad \text{for } 0 \leq x^* < \ell^* = \tau_o^* \quad (21 \text{ a})$$

$$= x^* - \tau_o^* \quad \text{for } \tau_o^* = \ell^* < x^* \leq 1 \quad (21 \text{ b})$$

For the case of $\pi^* > 1/\mu_{wL}$, Equations (18) and (19) must be modified to be

$$\text{at } x^* = 0, \quad \sigma_{xx}^* = \sigma_o^* = 0 \quad (22)$$

$$\tau_{xy}^* = \tau_o^* = 0$$

$$\text{at } x^* = 1, \quad \sigma_{xx}^* = \sigma_L^* = \pi^* \quad (23)$$

$$\tau_{xy}^* = \tau_L^* = 1$$

Note Equation (23) was obtained directly from Equations (22), (8), and (9).

The solutions to Equations (6) and (7) corresponding to the boundary conditions of Equations (22) and (23) give the stress distribution for $\pi^* > 1/\mu_{wL}$, or

$$\sigma_{xx}^* = \pi^* x^* \quad (24)$$

$$\tau_{xy}^* = x^* \quad (25)$$

The stress distributions are illustrated in Fig. 2.

Solid Velocity Profile

The heuristic argument used in obtaining the solid-velocity profile expressions can be stated as follows: Consider an imaginary plane (normal to the x-coordinate) in the solid phase. The frictional stress between particles on either side of the plane, τ_{f_i} (or $\tau_{f_i}^*$ in dimensionless form), can be related to the compressible stress at this point by the expression

$$\tau_{f_i}^* = \mu_i \sigma_{xx}^* \quad (26)$$

It is obvious that there can be no slip (i.e., relative motion) between particles on this plane if the frictional stress is greater than the shear stress at the plane, or

$$\tau_{f_i}^* - \tau_{xy}^* > 0 \quad (27 \text{ a})$$

Conversely, slip can be expected if

$$\tau_{f_i}^* - \tau_{xy}^* \leq 0 \quad (27 \text{ b})$$

Since the shear and frictional stresses are dependent upon μ_{w_o} , μ_{w_L} , and μ_i as well as upon the applied pressure gradient along the x-direction, π^* , velocity profiles of a few specific cases which are believed to be representative of those encountered in practical applications of cross-flow moving-bed filtration will be considered as detailed below.

Case (A) $\mu_{w_o} = \mu_{w_L} = \mu_w, \quad \mu_i > \mu_w, \quad \pi^* \leq 1/\mu_w,$

The stresses σ_{xx}^* and τ_{xy}^* given by Equations (20), (21 a), (21 b) and (18 b) become

$$\sigma_{xx}^* = \frac{1}{2\mu_w} + (x^* - \frac{1}{2}) \pi^* \quad (28)$$

$$\tau_{xy}^* = \frac{1}{2} (1 - \mu_w \pi^*) - x^* \quad 0 < x^* < \frac{1}{2} (1 - \mu_w \pi^*) \quad (29 \text{ a})$$

$$= x^* - \frac{1}{2} (1 - \mu_w \pi^*) \quad \frac{1}{2} (1 - \mu_w \pi^*) < x^* < 1 \quad (29 \text{ b})$$

The frictional stress can be found from Equations (28) and (26), or

$$\tau_{f_i}^* = \frac{\mu_i}{2\mu_w} + \mu_i \left(x^* - \frac{1}{2}\right) \pi^* \quad (30)$$

The quantity $\tau_{f_i}^* - \tau_{xy}^*$ is found to be

$$\tau_{f_i}^* - \tau_{xy}^* = (1 + \mu_i \pi^*) x^* + \frac{\mu_i - \mu_w}{2} \left(\frac{1}{\mu_w} - \pi^*\right) \quad \text{for } 0 < x^* < \frac{1}{2} (1 - \mu_w \pi^*) \quad (31 \text{ a})$$

$$= \left(\frac{\mu_i}{\mu_w} - 1\right) \left[x^* - \frac{1}{2} (1 - \mu_w \pi^*)\right] + \left(\frac{1}{\mu_w} - \pi^*\right) (1 - x^*) \mu_i$$

$$\text{for } \frac{1}{2} (1 - \mu_w \pi^*) < x^* < 1 \quad (31 \text{ b})$$

It is simple to show that every term of the above equations is positive.

Consequently, $\tau_{f_i}^* - \tau_{xy}^* > 0$, and there can be no slip between the granular particles.

Next consider the shear stresses and the frictional stresses at the two walls. From Equations (18 b), (19 b) and (29), one has

$$\mu_w \sigma_o^* - \tau_o^* = 0 \quad (32 \text{ a})$$

$$\mu_w \sigma_L^* - \tau_L^* = 0 \quad (32 \text{ b})$$

In other words, there is slip between the granular particles and the wall.

Thus, one would expect the solid to move as a plug, or

$$v^* = \frac{v_s}{\bar{v}_s} = 1 \quad \text{for} \quad u \leq x^* \leq 1 \quad (33)$$

where v_s and $\bar{v}_s$ denote, respectively, the solid velocity and its average value.

$$\text{Case (B)} \quad \mu_{w_o} = \mu_{u_L} = \mu_w \quad \mu_i > \mu_w \quad \pi^* \geq \frac{1}{\mu_w} > \frac{1}{\mu_i}$$

The quantity $\tau_{f_i}^* - \tau_{xy}^*$ can be obtained from Equations (24), (25) and (26), or

$$\tau_{f_i}^* - \tau_{xy}^* = \mu_i \pi^* x^* - x^* = (\mu_i \pi^* - 1) x^* \quad 0 \leq x^* \leq 1 \quad (34)$$

The above quantity is positive since $\pi^* > \frac{1}{\mu_i}$. So, there is no particle-particle slip in the solid phase.

The comparison between the shear stress and the frictional stress at the boundary [namely, Equations (22), (23), and (25)] yields

$$\mu_w \sigma_o^* - \tau_o^* = 0 \quad (35 \text{ a})$$

$$\mu_w \sigma_L^* - \tau_L^* = \mu_w \pi^* - 1 > 0 \quad (35 \text{ b})$$

Equation (35 b) suggests that there is no particle-wall slip at the downstream boundary. Furthermore, since there is no solid-velocity gradient throughout the bed, the solid phase must be at rest. The velocity profiles corresponding to cases (A) and (B) are shown in Fig. 3.

Case (C) $\mu_{w_o} = \mu_{w_L} = \mu_w$, $\mu_w > \mu_i$, $\pi \leq 1/\mu_w \leq 1/\mu_i$

The stress distribution is the same as for case (A), and the frictional stress is the same as in Equation (30), or

$$\sigma_{xx}^* = \frac{1}{2\mu_w} + (x^* - \frac{1}{2}) \pi^*$$

$$\tau_{xy}^* = \frac{1}{2} (1 - \mu_w \pi^*) - x^* \quad 0 \leq x^* \leq \frac{1}{2} (1 - \mu_w \pi^*)$$

$$= x^* - \frac{1}{2} (1 - \mu_w \pi^*) \quad \frac{1}{2} (1 - \mu_w \pi^*) < x^* \leq 1$$

$$\tau_{f_i}^* = \frac{\mu_i}{2\mu_w} + (x^* - \frac{1}{2}) \mu_i \pi^*$$

At the boundary $x = 0$, one has

$$\tau_o^* = \frac{1}{2} (1 - \mu_w \pi^*)$$

$$\mu_w \sigma_o^* = \frac{\mu_i}{2\mu_w} (1 - \mu_w \pi^*)$$

Similarly, at $x = 1$ ($x^* = 1$), one has

$$\tau_L^* = \frac{1}{2} (1 + \mu_w \pi^*)$$

$$\mu_w \sigma_L^* = \frac{\mu_i}{2\mu_w} (1 + \mu_w \pi^*)$$

With $\mu_w > \mu_i$, both $(\mu_w \sigma_o^* - \tau_o^*)$ and $(\mu_w \sigma_L^* - \tau_L^*)$ are negative. Equations (29 a) and (29 b) show that τ_{xy}^* decreases away from either boundary and vanishes at $x^* = \frac{1}{2} (1 - \mu_w \pi^*)$ and $\tau_{f_i}^*$ is always positive [see Equation (30)].

Consequently, $\tau_{f_i}^* - \tau_{xy}^*$ becomes positive over part of the interval $0 \leq x^* \leq 1$, as shown in Fig. 4. By simple algebraic manipulation, one can easily show that the positions τ_{xy}^* and $\tau_{f_i}^*$ are equal in magnitude, $x_{c_1}^*$ and $x_{c_2}^*$, are

$$x_{c_1}^* = \frac{(\mu_w - \mu_i)(1/\mu_w - \pi^*)}{2(1 + \mu_i \pi^*)} \quad (36 a)$$

$$x_{c_2}^* = \frac{(\mu_w + \mu_i)(1/\mu_w - \pi^*)}{2(1 - \mu_i \pi^*)} \quad (36 b)$$

In other words, $\tau_{f_i}^* - \tau_{xy}^* > 0$ for $0 \leq x^* < x_{c_1}^*$ and $x_{c_2}^* < x^* \leq 1$, and $\tau_{f_i}^* - \tau_{xy}^* < 0$ for $x_{c_1}^* < x^* < x_{c_2}^*$. One may, therefore, conclude that there is no inter-particle slip in the core, but slip does exist in regions near both the upstream and downstream boundaries. If the velocity profile near the boundary can be approximated as linear, then the velocity profile can be described as

$$\begin{aligned} v^* &= v_o^* + (v_c^* - v_o^*) \frac{x^*}{x_{c_1}^*} & 0 < x^* < x_{c_1}^* \\ v^* &= v_c^* & x_{c_1}^* < x^* < x_{c_2}^* \\ v^* &= v_L^* + (v_c^* - v_L^*) \frac{1 - x^*}{1 - x_{c_2}^*} & x_{c_2}^* < x^* < 1 \end{aligned} \quad (37)$$

where v_o^* , v_c^* , and v_L^* are the velocities at the upstream boundary, in the core, and at the downstream boundary. v_c^* can be expressed in terms v_o^* and v_L^* from material balance, or

$$\frac{1}{2} x_{c1}^* (v_o^* + v_c^*) + v_c^* (x_{c2}^* - x_{c1}^*) + \frac{1 - x_{c2}^*}{2} (v_c^* + v_L^*) = 1$$

Solving for v_c^* , one has

$$v_c^* = \frac{2 - v_o^* x_{c1}^* - (1 - x_{c2}^*) v_L^*}{1 + (x_{c2}^* - x_{c1}^*)} \quad (38)$$

The present analysis does not permit a direct evaluation of v_o^* and v_L^* . However, when the gas flow rate is sufficiently high and when granular particles verge on being pinned on the downstream boundary, it is easy to show that the shear stress at $x^* = 1$, τ_L^* , becomes unity. Based on dimensional reasoning, one may argue that

$$v_L^* = 1 - \tau_{xy}^* \Big|_L = \frac{1}{2} (1 + \mu_w \pi^*) \quad (39 a)$$

$$v_o^* = 1 - \tau_{xy}^* \Big|_o = \frac{1}{2} (1 - \mu_w \pi^*) \quad (39 b)$$

The velocity profile is shown in Fig. 4.

Case (D) $\mu_w > \mu_i$, $1/\mu_w < \pi^* < 1/\mu_i$

From Equations (22), (23), (24), (25), and (26), one has

$$\sigma_{xx}^* = \pi^* x^*$$

$$\tau_{xy}^* = x^*$$

$$\tau_{fi}^* = \mu_i \sigma_{xx}^* = \mu_i \pi^* x^*$$

$$\mu_w \sigma_L^* = \mu_w \pi^*$$

Since $1/\mu_w < \pi^* < 1/\mu_i$, τ_{xy}^* is greater than τ_{fi}^* over $0 < x^* < 1$, but $\mu_w \sigma_0^* - \tau_0^* = 0$ (at $x^* = 0$) and $\mu_w \sigma_L^* - \tau_L^* = \mu_w \pi^* - 1$ (at $x^* = 1$) is positive. Accordingly, the solid phase velocity must vanish at $x^* = 1$, but there is inter-particle slip throughout the solid phase. If, as before, the profile is assumed to be linear, then v^* is given as

$$v^* = 2 (1 - x^*) \quad (40)$$

Fig. 4 gives the shear stress and velocity distribution of this case.

Case (E) $\mu_w > \mu_i$, $\pi^* > 1/\mu_i > 1/\mu_w$

The values of σ_{xx}^* , τ_{xy}^* , τ_{fi}^* , $\mu_w \sigma_0^*$, and $\mu_w \sigma_L^*$ are the same as those of Case (D). However, since $\pi^* > 1/\mu_i$, τ_{fi}^* is greater than τ_{xy}^* . Consequently, the solid phase is at rest.

Comparisons with Experimental Data

Mizukami et al. (1984, 1985) reported solids flow data on cross-flow moving-bed filters under two sets of conditions. Their data and predictions from the analysis reported above were compared in order to validate the present work. The experiments were conducted using a moving-bed filter of 10 x 20 x 60 cm (L x W x H, where L is the distance gas flows, W, the column width; and L, its height along the direction of solid flow). Silica gel particles (with diameter between 1.2 to 2.5 mm) were the filter medium. The solids velocity was determined by monitoring the distribution of tracer particles (white-

colored spheres). In addition, the investigators obtained the angle of friction data (with the use of the Jenike cell), from which μ_w and μ_i can be obtained. Details of the experimental conditions are given in Table 1.

The first set of data was obtained with the walls of the column made of perforated plates. The friction coefficient, μ_w , was found to be 0.65 (or $1/\mu_w = 1.54$). Since the internal friction coefficient, μ_i , was estimated to be 0.87, or $\mu_i > \mu_w$, the solid-velocity profile is expected to be flat if the dimensionless pressure gradient, $\pi^* = (-dp/dx)/\gamma$, is less than $1/\mu_w = 1.54$ [(in other words, case (A)); there is no motion if π^* is greater than 1.54.

Experiments and predictions are compared in Fig. 5. As expected, the solid flow data with no air flow gave essentially a flat profile.

The solid-velocity profile was found to be flat at air velocity $u = 1.4$ m/s (corresponding to $\pi^* = 1.11$), and the solid was found at rest at $u = 2.2$ m/s (corresponding to $\pi^* = 1.58$). In both cases, the observations agree with predictions. On the other hand, at $u = 1.8$ m/s (or $\pi^* = 1.32$), part of the solid (that near the downstream side) became stalled, which was not predicted.

For the second set of results, the walls of the moving bed were made of steel wire screening which has a higher value of μ_w than that of the first case, $\mu_w = 0.98$. For $\mu_w > \mu_i$, the solid-velocity profiles are given by Equations (37) or (40), depending upon the relative magnitude of $1/\mu_i$ and π^* .

The comparisons between predictions and experiments are shown in Fig. 6. For $u \leq 1.0$ m/s, with $\pi^* < 1/\mu_w = 1.02$, the predicted velocity profile agrees well with experiment. For air velocity $u = 1.70$ m/s, which gives a pressure gradient greater than $1/\mu_w$, the solid was supposed to have a linear velocity

profile with vanishing velocity at the downstream side. The solid was, however, found to stall over a distance from the downstream wall.

From these comparisons, it is clear that the present analysis, in spite of its extreme simplicity, is essentially accurate in predicting the solid-flow behavior in cross-flow moving beds. As the solid-flow pattern is of crucial importance to the performance of cross-flow granular moving-bed filters, the results of this work should prove useful to the design and modelling of granular moving-bed filtration systems.

Table 1

Experimental Conditions Used by Mizukami et al. (1984, 85)

$$\gamma = 1.53 \text{ grams/cm}^3$$

$$d_p = 1.2 \sim 2.5 \text{ mm}$$

$$\mu_i = 0.87$$

$$\mu_w = 0.98 \text{ (wire screens)}$$

$$\mu_w = 0.65 \text{ (perforated plates)}$$

$$u_s = 0 - 2.2 \text{ m/s}$$

$$(-dp/dx) = 0 - 2.42 \text{ cm Aq/cm}$$

$$L = 10 \text{ cm}$$

FIGURE 1 SCHEMATIC DIAGRAM OF CROSSFLOW MOVING BED

FIGURE 2 STRESS DISTRIBUTION DIAGRAMS

FIGURE 3 SOLID VELOCITY PROFILES FOR THE CASES

$$\mu_{w0} = \mu_{wL} = \mu_w, \mu_i > \mu_w$$

$$\pi^* \leq \frac{1}{\mu_w} < \frac{1}{\mu_i}$$

$$\frac{1}{\mu_w} < \pi^* < \frac{1}{\mu_i}$$

FIGURE 4 SOLID VELOCITY PROFILES FOR $\mu_{w0} = \mu_{wL} = \mu_w$ and $\mu_w > \mu_i$

FIGURE 5 COMPARISON OF PREDICTIONS WITH EXPERIMENTS FOR THE CASES
 $\mu_i > \mu_w$ $\mu_i = 0.87, \mu_w = 0.65$

FIGURE 6 COMPARISON OF PREDICTIONS WITH EXPERIMENTS FOR THE CASE
 $\mu_w > \mu_i$, $\mu_w = 0.98, \mu_i = 0.87$

Notation

k	Rankin constant [Equation (12)]
L	Filter medium depth
ℓ^*	quantity defined by Equation (21.a)
P	pressure
v_s	solid velocity
$\bar{v}_s$	average solid velocity
v^*	defined as $v_s / \bar{v}_s$
v_o^*, v_L^*	dimensionless value of v^* at $x^* = 0$ and $x^* = 1$
v_c^*	dimensionless value of v^* in the solid case
x, y	coordinates
x^*, y^*	dimensionless coordinate
x_{c1}^*, x_{c2}^*	defined by Equations (36.a) and (36.b)

Greek Letters

π^*	defined as $(-\partial p / \partial x) / \gamma$
γ	solid weight
σ_{xx}, σ_{yy}	compressive stress components
σ_o, σ_L	values of σ_{xx} at $x = 0$ and $x = L$, respectively
$\sigma_{xx}^*, \sigma_{yy}^*$	values of σ_{xx} at $x = 0$ and $x = L$, respectively
σ_o^*, σ_L^*	dimensionless value of σ_o and σ_L
τ_{xy}, τ_{yx}	shear stress components
τ_o, τ_L	values of τ_{xy} at $x = 0$ and $x = L$, respectively
τ_{xy}^*	dimensionless value of τ_{xy}
τ_o^*, τ_L^*	dimensionless values of τ_o and τ_L
τ_{fi}	frictional stress
μ_i	friction coefficient
μ_{wo}, μ_{wL}	wall friction coefficient at $x = 0$ and $x = L$
μ_{wT}	defined as $\mu_{wo} + \mu_{wL}$

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Part V.

Aerosol Deposition in Granular Moving Bed

Introduction

Among various particulate emission-control processes, granular filtration is known for its relatively high collection efficiency for particles of widely ranging size and for its simplicity in operation. Other advantages of granular filtration include the abundance of granular substances which can be used as filter grains and the fact that many available granular substances are immune to high-temperature, high-pressure, and/or chemically corrosive environments. As a result, granular filtration has been widely used in gas cleaning and dedusting, with applications in practically every industrial segment (Juvinall et al., 1970).

Granular filtration is commonly conducted in the fixed-bed mode. Fixed-bed filtration, however, is inherently non-steady in nature. Initially free of any deposited matter, a filter bed becomes increasingly clogged with deposited particles during the course of filtration. Clogging, in turn, increases the pressure drop necessary to maintain a constant gas throughput. Ultimately, the pressure drop across the filter becomes so excessive that one must stop the filtration process to replace or regenerate the filter medium.

One alternative for overcoming the disadvantages arising from fixed-bed clogging is to use a moving bed. In one such approach, cross-flow moving-bed filtration, the flow of the gas suspension is transverse to the flow of the filter grains. This arrangement, which combines the advantages of granular filtration and the benefits of continuous operation, has been seriously considered for use in hot gas cleaning and, indeed, is the basis for a patent granted more than sixty years ago (Klarding, 1921).

Despite its potential, cross-flow moving-bed granular filtration needs to be examined much more rigorously before its true value can be assessed. The present work represents the first phase of a systematic study in cross-flow moving-bed granular filtration aimed at developing a comprehensive framework that accurately describes the essential features of the process. The specific aims of this work are:

- (1) to determine the extent of particle collection achieved and the effect of various operating variables on particle collection;
- (2) to examine the possible operational complexities arising from the interaction between solids flow and gas flow;
- (3) to delineate the essential difference between fixed-bed and moving-bed filtration.

Experimental Work

Apparatus: A schematic diagram of the experimental apparatus is shown in Fig. 1. The moving bed was made of a brass sheet (1/16 inch in thickness) with three sections (entry section, filter section, and discharge section) flanged together. The entry section began with a hopper (for solids storage) which had a cross section of 127 mm by 82.6 mm and a height of 470 mm. The hopper tapered into a duct of rectangular cross section measuring 20.3 mm by 82.6 mm by 308 mm high. The discharge section was 635 mm high with a cross section similar to that of the entry section.

The filter section was composed of a solids flow duct (280 mm high) and a gas flow duct (140 mm long). Both ducts had identical cross sections (20.3 mm by 82.6 mm). They were joined perpendicularly at their midsections. To prevent any loss of filter grains into the gas flow duct, mesh wire screens

(with 150 μm openings) were placed on the two sides of the solids flow duct to which the gas flow duct was joined. The gas flow duct was preceded by a divergent flow section and followed by a discharge duct. Provisions were made to sample both the influent and effluent gas streams and to measure the pressure drop between the upstream and downstream side.

Downward movement of the filter grains was maintained at a velocity of 0.2 to 2 mm/s with a DC motor (1/6 HP and 1750 rpm, Boston Gear Co.) and a worm gear reducer (Model 001, Boston Gear Co.) controlled by a DC motor controller. Details of the moving bed are shown in Fig. 2.

The apparatus used for preparing the test suspensions is shown in Fig. 1. All the suspensions were prepared using a collision atomizer (BGI Inc., Waltham, MA) and monodispersed polyvinyltoluene latex particles (Dow Chemical Co., Indianapolis, IN). Bottled nitrogen gas at 35 psig or 70 psig was passed through the atomizer's aspirators to create a spray. The spray was first passed through a heating tube (namely, a glass tube of 5/8 in. i.d. with heating tape wrapped around the tube) in order to vaporize the water on the particle surface. The spray was then sent through a diffusion dryer to remove the water vapor before entering a neutralizer (Model 3012, TSI, St. Paul, MN), where the electrostatic charge was reduced to a minimum. The test suspension entered the filter through the upstream side of the gas entry section, which, as shown in Fig. 2, was placed ahead of the gas duct of the cross-flow filter.

The experimental work measured the concentrations of the influent and effluent gas suspensions and the pressure drop associated with the gas flow. A particle counter (Model APS 3300 TSI, St. Paul, MN) was used to give the number concentrations of the monodispersed latex particles in the gas streams, and a manometer provided the pressure-drop readings.

Experimental Systems

The filter grains used were nearly mono-sized glass spheres with average diameters of 280 μm and 525 μm and magnetite particles with an average diameter of 367 μm . Under microscopic examination, the glass spheres were found to be nearly uniform in geometry and with smooth surfaces (Chiang and Tien, 1985). The latex particles used had a diameter of 2.02 μm .

Experimental Conditions

The conditions used in carrying out the experiments were:

Filter Grain Diameter	280 μm , 367 μm and 525 μm
Particle Diameter	2.02 μm
Filter Width	82.6 mm
Solids Velocity	up to 16 cm/min
Gas Velocity	> 5 cm/s
Filter Grain Residence Time	> 7.5 s
Time Duration of Measurement	$\leq 62/v$ min when v is the solids velocity in cm/min

Results

A substantial number (over thirty) of experimental runs were made to obtain data on filter performance (c_{eff} and p vs. time). The salient features of the results are given below.

Possible Interaction Between Gas and Solid Flow. For proper operation of cross-flow moving-bed filters, transverse gas flow must not impede solid flow. Otherwise, as shown by previous investigators (Bridgwater, 1982; Ginestra and Jackson, 1985), for high gas velocities, granular particles constituting the filter medium may be pinned against the downstream boundary of the moving bed. Under extreme conditions, solid flow may cease totally.

For the present work, it was important to take measurements when such gas-solid interaction was absent. To insure that the necessary experimental conditions were maintained, gas pressure-drop data over a period of time and corresponding to two different solid velocities were obtained. These data are shown in Fig. 3. The fact that the pressure drop was independent of the solid velocity and time indicates that the conditions used for conducting the measurements were indeed proper.

Particle Collection. Filter performance is usually described by the so-called total collection efficiency, E , defined as

$$E = \frac{c_{in} - c_{eff}}{c_{in}} \quad (1)$$

where c_{in} and c_{eff} denote, respectively, the influent and effluent concentrations.

Typical results which demonstrate the extent of particle collection achieved in the experimental moving-bed filters are shown in Figs. 4 and 5. In both figures, the results are presented in the form of $1 - E$ or P , [the penetration, defined as $c_{\text{eff}}/c_{\text{in}}$] vs. the dimensionless time, t/θ , where θ is the average residence time of the filter grain, defined as

$$\theta = H/v \quad (2)$$

where v is the solid velocity and H is the height of the moving-bed filter. Note that H is not the same as the depth of the filter medium.

The results shown in Fig. 4 were obtained at two gas velocities ($u = 5.0$ and 10.0 cm/sec) and at a fixed solid velocity ($v = 4$ cm/min). The extent of penetration observed for the case of $u = 10$ cm/s was lower than the penetration for the case of $u = 5$ cm/s, which is consistent with the expectation that particle collection attributable to inertial impaction increases with the increase of gas velocity. The effect of solid velocity on particle collection was found to be negligible, as shown by the results displayed in Fig. 5.

Both the results showed some scattering in the values of $1-E$. However, they do not exhibit any definite trend as was the case for all of the results collected. Equally important is the fact that the pressure drop associated with gas flow was essentially constant. This behavior is in marked contrast to that observed in fixed-bed granular filtration. For fixed-bed filtration, both the collection efficiency of a filter and the pressure drop necessary to maintain a given flow rate through the filter increase with time because of the progressive clogging taking place in the bed. In a moving-bed filter,

the filter grains are renewed continuously. Furthermore, since the filter grains are in motion, it becomes more difficult for deposits to form on adjoining or neighboring filter grains. Consequently, the effect of deposition on particle collection can be expected to be small. As an approximation, one may assume that the unit collector efficiency of the filter grains along the gas-flow path is constant and equal to the initial unit collector efficiency, η_0 . The relationship between η_0 and E is given as (Pendse and Tien, 1982)

$$\eta_0 = \frac{dg}{L} \left[\frac{\pi}{6(1-\epsilon)} \right]^{1/3} \ln(1 - E) \quad (3)$$

where dg , L , and ϵ are the grain diameter, the depth of filter medium (along the direction of gas flow), and the bed porosity, respectively.

The unit collector efficiency provides a meaningful indicator of the grain's intrinsic ability to capture particles. For example, η_0 is independent of the filter depth, while E is strongly influenced by it. Because of this consideration, values of η_0 corresponding to those shown in Figs. 4 and 5 were obtained and presented in Figs. 6 and 7.

A number of empirical expressions have been proposed and can be used to estimate the initial unit collector efficiency of granular fixed-beds. The correlation proposed by Yoshida and Tien (1985) is given as

$$\eta_0/B = 100 N_{St}^2 + 0.19 \left(4 - \frac{4N_R}{d_c^*} + \frac{N_R^2}{d_c^{*2}} \right)^{1/2} \frac{N_R^{1.041}}{d_c^*} \quad N_R \leq 0.002 \quad (4a)$$

$$\eta_0/B = N_{St} + 0.48 \left(4 - \frac{4N_R}{d_c^*} + \frac{N_R^2}{d_c^{*2}} \right)^{1/2} \frac{N_R^{1.041}}{d_c^*} \quad N_R \geq 0.002 \quad (4b)$$

and

$$B = 7 - 6 \exp [-0.0065 N_{Re_s}] \quad (4c)$$

where

$$N_R = d_p / d_g \quad (5a)$$

$$N_{Re_s} = d_g u \rho / \mu \quad (5b)$$

$$N_{St} = d_p^2 u \cdot \rho_s C_s / 9 \mu d_g \quad (5c)$$

where d_p is the particle diameter; u , the gas velocity (superficial); and ρ , μ , and ρ_s , the gas density, gas viscosity, and particle density, respectively. C_s is the Cunningham correction factor, and d_c^* is the dimensionless average pore constriction diameter, which can be taken to be 0.35 for beds packed with nearly monosized grains.

The values of η_o obtained experimentally and those estimated from Equations (4a) through (4c) were compared, with the results also shown in Figs. 6 and 7. The comparisons clearly indicate that the extent of particle collection achieved in moving-bed filters is comparable to that of clean fixed-bed filters. In this connection, one should bear in mind that in estimating the unit collection efficiency, empirical correlations [including that given by Equations (4a) through (4c)], are usually accurate within a factor of two.

Another important aspect in moving-bed granular filtration is the triboelectric effect resulting from the relative motion between filter grains themselves and that between filter grains and the filter bed boundary. Tardos et al.

(1983) found in fluidized filter studies that the triboelectric effect significantly enhances particle collection.

The presence of the triboelectric effect was observed in several ways. During the initial phase of the study, when trial experiments were performed, it was rather difficult to obtain reproducible results. It was then found that discrepancies in collection efficiency obtained under seemingly identical conditions could largely be attributed to whether or not filter grains were in motion prior to the data collection. That presence or absence of movement, determines, of course, the presence of the triboelectric effect.

One series of experiments to demonstrate the presence of the triboelectric effect was performed as follows: The moving bed was first set in motion for five minutes at a downward velocity of 4 cm/s; it was then stopped. The bed was then operated as a fixed-bed filter, with the total collection efficiency determined over a period of time. A set of data obtained is shown in Fig. 8, in which the total collection efficiency is plotted as a function of time.

The $(1-E)$ vs. time shown in Fig. 8 exhibits a decrease in E (or an increase in penetration, $1-E$) with time. It is also shown that a significant difference in E occurs before and after particle motion. The results also indicate that after sufficient time had elapsed after the cessation of particle motion, E becomes comparable to those values obtained before particle motion. The behavior observed over all agrees with the facts that (a) inter-particle friction and particle-wall friction caused by particle motion generate electrostatic charges; (b) the charge is greatest immediately after motion stops but declines with time; and (c) electrostatic charge enhances particle collection.

A further evidence of the presence of the triboelectric effect can be seen from the data shown in Fig. 9. The data for Fig. 9 were compiled from measurements collected in the following manner. Initially, the filter was empty, and filter grains were poured into it. Immediately after the bed was filled, filtration (in the fixed-bed mode) measurements were made and collection efficiencies determined. The bed was then allowed to stand for 24 hours before another filtration measurement was taken. Immediately after the second measurement, the bed was set in motion for 20 minutes. Another fixed-bed measurement was taken. The bed was then allowed to stand for 24 hours, and the procedures were repeated.

The results shown in Fig. 9 indicate that the collection efficiencies achieved immediately after the filter grains had been set in motion were consistently greater than those measured after the bed had been at rest for 24 hours. The contribution of electrostatic charge is obvious.

Conclusions

The experimental work carried out in this study demonstrates clearly that under conditions where solid flow rate is kept moderately low (≤ 6 cm/min), the solid motion does not interfere with gas flow in the operation of cross-flow moving-bed filters. Both the collection efficiency and pressure drop do not exhibit time-dependent behavior, and the extent of particle collection can be estimated using the simple collection efficiency correlation recently proposed by Yoshida and Tien.

Nomenclature

c_{in}	influent concentration (numbers of particles/cm ³)
c_{eff}	effluent concentration (numbers of particles/cm ³)
C_s	Cunningham's correction factor (-)
d_g	filter grain diameter (mm)
d_p	aerosol particle diameter (μm)
E	total collection efficiency (-)
H	height of filter (mm)
L	filter width (mm)
N_R	dimensionless group defined by Equation (5a) (-)
N_{Res}	dimensionless group defined by Equation (5b) (-)
N_{St}	dimensionless group defined by Equation (5c) (-)
t	time (s)
u	gas velocity (cm/s)
v	solid velocity (mm/min)

Greek Letters

ϵ_o	bed porosity (-)
ρ	gas density (kg/m ³)
ρ_s	particle density (kg/m ³)
η_o	unit collector efficiency (-)
θ	defined as H/v (min)
μ	gas viscosity Pa-s

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FIG. 1 GENERATION OF AEROSOL SUSPENSION AND ITS FILTRATION

FIG. 2 SCHEMATIC DIAGRAM OF MOVING-BED FILTER

FIG. 3 PRESSURE DROP IN THE DIRECTION OF GAS FLOW

FIG. 4 AEROSOL PENETRATION AT TWO GAS VELOCITIES

FIG. 5 AEROSOL PENETRATION AT TWO FILTER GRAIN VELOCITIES

FIG. 6 UNIT COLLECTION EFFICIENCY OF MOVING-BED FILTRATION - EFFECT OF GAS VELOCITY

FIG. 7 UNIT COLLECTION EFFICIENCY OF MOVING-BED FILTRATION-EFFECT OF FILTER GRAIN VELOCITY

FIG. 8 AEROSOL PENETRATION CHANGE WITH TIME

FIG. 9 EXPERIMENTAL EVIDENCE ON THE GENERATION AND DISSIPATION OF ELECTRICAL CHARGES

Part VI.

Aerosol Deposition in a Magnetically Stabilized Fluidized Bed

Introduction

The use of an externally applied magnetic force field to maintain homogeneous gas fluidization was established conclusively by Rosensweig (1978, 1979) sometime ago. With the maintenance of homogeneous fluidization and the elimination of gas bubbles, a stable fluidized bed has the combined advantages of both conventional fluidized bed (continuous operation) and fixed bed (high efficiency). A natural application of magnetically stabilized fluidized beds (MSB) is their utilization used in aerosol filtration.

Albert and Tien (1985) conducted experiments on aerosol filtration in a magnetically stabilized fluidized bed. Their experimental work used magnetite granules as filter medium and aerosols prepared by dispersing talcum powder in air. The experimental variables included the magnetic granule size, bed height, applied magnetic field strength and air velocity. Their results can be summarized as:

1. The total collection efficiency increased with bed stability, (i.e., the efficiency increased with increasing magnetic field strength and decreasing air velocity);
2. The total collection efficiency decreased with time;
3. Beds of small granules channeled more readily than beds of large granules;
4. Significant particle re-entrainment existed in an MSB;
5. Packed bed efficiencies were possible in an MSB;
6. The filter coefficients showed packed bed tendencies.

A subsequent study by Warrior and Tien (1986) continued the earlier work of Albert and Tien and focussed on the possible reentrainment of collected aerosols and the behavior of an MSB after significant deposition occurred in the bed. Their work yielded the following results:

1. The collection efficiencies were much higher in an MSB compared to a simple fluidized bed;
2. There was no significant re-entrainment in an MSB;
3. Gas bypassing and bubbling were brought about in stable beds by reduced inter-particle magnetic forces resulting from surface deposits;
4. In general, long beds remained more stable after long hours of operation as a filter;
5. The initial filter coefficients decreased with increasing specific deposit due to instabilities caused by bed loading of aerosols.

Warrior and Tien's work raised certain potentially significant questions. From the point of view of effective particulate removal, in practical applications, the instability (and its manifested decline in collector efficiency) observed in Warriors and Tien's work must be avoided. For this purpose, an understanding of the filtration performance in the various MSB structures is required. It is therefore necessary to examine the MSB structure and relate filtration performance with the various types of bed structures. The specific objectives of the present experimental work are:

1. To gain a better understanding of the MSB structures.
2. To develop a stability or phase diagram from pressure drop/flow rate experiments.
3. To observe the behavior of an MSB operating near or in the marginal stability zone.
4. To determine the stability of an MSB with large amounts of aerosol deposition.

Experimental Work

Apparatus. A schematic diagram of the experimental apparatus used in this work is shown in Fig. 1. The apparatus is similar to that used previously

by Albert and Tien (1985) and Warrior and Tien (1986) but with certain modifications (principally the solenoids for stabilizing the fluidized filter). The apparatus consisted of the following main components; the air supply system, the aerosol generator, the experimental filter and the sampling system.

Air was supplied from a compressor (Dayton Speedaire Model 3Z582 single stage) and passed through a series of prefilters, filters and dryers in order to remove impurities and moisture. The air stream was then split into two parts, with one part used to operate an aerosol generator (TSI Model 3400 Fluidized Bed Aerosol Generator), for preparing test aerosols. The other part of the air stream was regulated by an air regulator (Perfecting Series R-13 Air Regulator) and used as a dilution stream to be mixed with the concentrated aerosol stream exiting from the aerosol generator.

The test aerosols entered into a 3 in. i.d. plexiglass column (mixing chamber) in order to remove the entrained particle agglomerates and heavy particles. The aerosols then entered the experimental filter consisting of a plexiglass column (5.08 cm i.d. and 71 cm height). The distributor of the fluidized filter was a 48 Tyler mesh screen with opening of 297 μm . Two pressure taps (one placed 10 cm below the distributor and the other 55 cm above the distributor) were inserted into the column walls. A disengaging chamber was placed on top of the filter. There were two sample ports for measuring the influent and effluent concentrations. The influent sample port was placed 14 cm below the distributor and the effluent sample port was located at the top of the disengaging chamber.

The test aerosols upon leaving the filter first passed through an absolute filter (to remove all the particles) and then a rotameter (Dwyer RMC-40-55V Rate Master Flowmeter) before discharging to the ambient.

To obtain the required magnetic stabilization effect, the experimental filter was surrounded by two solenoids which were fabricated by winding 1,000 times (approximately 560 meters) of 18 AWG magnet wire around plexiglass columns 15.2 cm in i.d. and 8 cm in height. Each coil has a resistance of 11.84 ohms. When connected in series, they could produce a maximum field intensity of 2.0 Oersteds.

Pressure drop across the experimental filter was measured by a Druck PDCR 10/L Pressure Transducer connected to a Druck Model DPI 203 Digital Pressure Indicator, the signal of which was sent to a chart recorder (Houston Instrument Omniscribe Chart Recorder). The influent and effluent concentrations were determined by weighing the amount of aerosol particles collected on the filter paper (Gelman glass-fiber filter paper) placed in the sample holder. The sample holder used and the sampling procedures were exactly the same as those used previously (Albert and Tien, 1985; Warrior and Tien, 1986).

System. The filter grains used were magnetite particles (NL Chemical Company, MacIntyre Division). The magnetite particles were sieved to 24-48 Tyler mesh size, washed and then dried before used in experiments. The average granular size was determined using a Dynascope Optical Beam Projector, Model 7A. The relevant magnetite properties are given in Table 1.

Talc powder [approximated by the formula $\text{Mg}_6(\text{Si}_2\text{O}_5)_4(\text{OH})_4$] was used to prepare the test aerosols. The average powder size determined by a TSI Aerodynamic Particle Analyzer was 1.6 μm .

Procedures. The procedure used to examine the bed structure and that used to conduct filtration measurement are described separately in the following:

The structure of the experimental filter under magnetic stabilization was examined by establishing the pressure drop-flow rate relationship. The

experiments were initiated by setting the current in the coils to attain the desired magnetic field strength. The air compressor was started, and the flow rate through the MSB was increased in increments of 10 to 20 ft³/hr. The pressure drop and the bed height were recorded at each flow rate.

The flow rate was increased until the bed bubbled continuously or until further increase was impossible because of the compressor's limitations. At this point, the flow rate was decreased in increments of 20 ft³/hr until air flow ceased. Again, the pressure drop and the bed height were recorded at each air flow setting. The bed was observed throughout the experiments.

To begin filtration experiments, the magnetic field was applied and the compressor was switched on allowing air to flow through the system. The aerosol generator flow rate was set and the purge line opened to prevent aerosols from entering the MSB. After the aerosol generator emitted a steady stream of powder, the purge line was closed and the aerosol-laden gas was allowed to mix with the dilution stream of air and to enter the MSB. Depending on the desired bed structure, the air flow rate through the MSB was set in either of the following ways:

1. The air flow was increased gradually to the desired setting, where it remained throughout the entire experiment;
2. The air flow was increased to a previously chosen value higher than the operating condition. Once this value was reached, the flow rate was decreased to the desired experimental setting.

The Gelman glass-fiber filter papers were placed in the sample holders, and the flow rates through each sampler were adjusted. The filter papers in the sample holders connected to the influent and effluent sample ports were changed every 30 minutes and were weighed on an analytical balance to determine the weight of powder collected. The pressure drop and the bed

height were recorded prior to changing the filter papers. Filtration experiments varied in duration from three to ten hours. The compressor was shut off at the end of the experiment, and the magnetite particles were removed from the MSB. To minimize the effect of aerosol collection by the walls of the column, the MSB was never cleaned.

The experimental variables studied included the gas velocity (up to $2.0 u_{mf}$ where u_{mf} is the minimum fluidization velocity), the bed height (8 cm in unfluidized state) and the applied magnetic field intensity (up to 15 Oersteds).

Structures of Magnetically Stabilized Fluidized Beds

Because of the structure of a fluidized filter and the collector efficiency of the filter are intimately connected, a better understanding of the structures of the MSB was considered essential to a proper interpretation of the filtration data. The results obtained from the series of pressure drop-flow rate measurements obtained at various applied magnetic field intensity can be summarized as follows:

Fluidization Without Magnetic Stabilization. As a basis of comparison, flow rate-pressure drop measurements of fluidized beds without magnetic stabilization were made. In Fig. 2, a typical set of such data is displayed. When the air flow rate increased from zero to the point of incipient fluidization, the pressure drop increased linearly with increasing gas velocity as the Carman-Kozeny equation predicts. At the onset of fluidization, (or $u = u_{mf}$ - 30 cm/s), the pressure drop equalled the weight of the magnetite granules per unit cross-sectional area of the bed, W/A , as dictated by the fluidization theory. Further, increasing the gas velocity resulted in bed bubbling.

The pressure drop in the bubbling regime declined to a constant value of approximately 0.8 W/A. Observation of the bed revealed unfluidized or dead regions. The decline in pressure drop and the presence of unfluidized regions, in all likelihood, were the result of bed channeling. Leva (1959) discussed the effects of channeling in a fluidized bed and attributed channeling to the following: 1) large granule density, 2) non-uniform gas distribution, or 3) non-spherical granules. All of these factors were present in the study.

When the air velocity was reduced, the pressure drop remained constant at 0.8 W/A until the bed ceased bubbling. At that point, the pressure drop followed a new path with decreasing air velocity. The pressure drop-gas velocity relationships obtained by decreasing air velocity was below that obtained from increasing air velocity. The bed height data during fluidization-defluidization as presented in Fig. 3 show that the bed height expanded 1 cm upon fluidization and remained expanded after defluidization. This bed expansion corresponded to an increase in porosity from 0.47 (the dumped-bed porosity) to 0.54 (the expanded-bed porosity).

Fluidization Under Magnetic Stabilization. Pressure drop-flow rate experiments were performed similarly to those described above except that a magnetic field was applied before starting the gas flow. The general trend observed, which Rosensweig (1979) also reported, was that higher magnetic field strengths provided increased bed stability, i.e., no bubbling.

Fig. 4 shows the pressure drop-flow rate behavior for an MSB with an applied field of 50 Oersteds. The path followed by the pressure drop when increasing the air velocity from 0 to u_{mf} was similar to that shown in Fig. 2. These results compare well with similar ones obtained by Rosensweig (1979), who stated that the behavior of the fixed bed region can be predicted by

the Ergun equation (a modified version of the Carman-Kozeny equation) and is not affected by an applied magnetic field.

The magnetic field caused a plateau to develop between the minimum fluidization velocity, u_{mf} , and the transition velocity, u_t . Without the applied magnetic field, the transition velocity coincided with the minimum fluidization velocity. With an applied magnetic field, a difference resulted between u_{mf} and u_t and this difference between u_{mf} and u_t increased with magnetic field strength (for example, compare Fig. 4 with 5). Rosensweig termed the region between u_{mf} and u_t as the stably fluidized regime.

When the air velocity reached u_t in the unmagnetized bed, the bed bubbled or spouted continuously; such activity did not occur in the magnetized bed. Upon reaching u_t , the magnetized bed bubbled briefly but returned to a metastable state after the granules formed a new structure (bed channels were usually formed during this time). As the air velocity was increased beyond u_t , the process of bubbling-rearrangement was repeated until constant bubbling occurred at higher velocities. In weaker magnetic fields, e.g., 50 Oersteds, continuous bubbling took place at velocities close to u_t . The structures were considered unstable after initial bubbling occurred followed by a decline in the pressure drop; a result of bed channeling which was similar to the unmagnetized bed. As with the unmagnetized bed, bed channeling caused the pressure drop after bubbling to decrease by over 20 percent.

An interesting phenomenon was observed in the magnetized beds upon defluidization. As shown in Fig. 5, the pressure drop did not retrace itself when the air velocity was first increased beyond the point of incipient fluidization and then decreased. Hysteresis was observed in the pressure drop-flow rate curves. Furthermore, the pressure drop declined linearly, and the bed height and the bed porosity remained constant while the air velocity was reduced.

Three experiments which further demonstrate the behavior described above are plotted in Fig. 6. In each run an applied magnetic field of 100 Oersteds was used. In two of the experiments, the air velocity was decreased from the stably fluidized regime, and in the third run, the air velocity was decreased from the bubbling regime. Hysteresis was prevalent in all three runs. The bed heights and bed porosities were unchanged upon defluidization; the pressure drop in all cases declined linearly when the air flow was reduced. The behavior exhibited by the defluidization MSB structures suggests that fixed beds were formed at air velocities in excess of u_{mf} .

To determine the MSB structure formed upon defluidization, the experimental pressure drop data were compared with values calculated from the Carmen-Kozeny equation. In making the comparison, bed porosities were calculated from bed expansion data. Table 2 compares the pressure drop data shown in Fig. 6 with the calculated results. Runs P5 and P6 compare well with the Carmen-Kozeny equation. In contrast, the agreement between experimental and theoretical is poor for run P3. The poor agreement can be attributed to channeling within the bed which formed after the bed bubbled and rearranged its structure. Because of gas bypassing, the resulting pressure drop data were much lower than those predicted by the Carmen-Kozeny equation. However, the agreement of runs P5 and P6 with the Carmen-Kozeny equation confirms that the structure obtained upon defluidization were fixed beds.

The above described phenomenon was also observed by Arnaldos et al. (1983). Two limiting cases concerning the rearrangement of granules were noted by Arnaldos et al. and are illustrated in Fig. 7. The first case suggested that the granules were loosely packed, and that the interstitial conduits were tortuous. The second case described the granules as being perfectly aligned along the lines of the magnetic field.

This study produced evidence supporting the theory proposed by Arnaldos et al. Bed height versus air velocity data are plotted in Fig. 8. As previously shown, the bed height for run P1, the unmagnetized case, expanded the approximately 1 cm upon fluidization-defluidization. With the applied magnetic field, the bed heights (from runs P3 and P4) expanded by approximately 2 cm after fluidization-defluidization. The bed remained in its expanded state because of the residual magnetic effects that existed after the air flow was shut off.

The Stability Diagram. Based on the data from the pressure drop-flow rate experiments, we constructed a stability diagram by plotting u_{mf} and u_t as a function of the applied magnetic field intensity. As shown in Fig. 9, the stability diagram, which is, in a large measure, similar to the work presented by Rosensweig (1979), exhibits these regions: unfluidized, stably fluidized, unstably fluidized and marginally stably fluidized. The magnetic field strength did not affect the minimum fluidization velocity but did cause the transition velocity to increase at higher magnetic field strengths.

In contrast to Rosensweig's analysis, the transition velocities were not always found to be sharp and reproducible. Colberg and Liu (1983) showed that non-spherical granules such as the magnetite grains used in the present study did not give reproducible transition velocities. To account for this effect, one may argue that there exists a zone of area of marginal stability in the stability diagram as opposed to the line between stably and unstably fluidized states presented by Rosensweig. The width of the marginal zone was estimated to be 10 cm/s, corresponding approximately to the difference among the experimentally determined transition velocities.

The transition zone in Fig. 9 was calculated by a least-squares fit using the mean of the transition velocities at each magnetic field strength. The resulting linear equation is

$$u_t = 0.71 \times H - 1.5 \quad (1)$$

The marginal stability line obtained from the least-squares fit was extended by ± 5 cm/s to account for the non-reproducible transition points.

Filtration Equipment

A number of filtration experiments were made under conditions listed in Table 3. The corresponding positions of these experiments in the stability diagram can be found readily by referring to the location number shown in Fig. 9 and those listed in Table 3.

For each experiment, the following variables were measured; the influent and effluent concentration, c_{in} , and c_{eff} , the pressure drop across the filter, p , and filter height at various times, H . For the concentration measurements, the amount of particles collected within the filter, M and the collection efficiency, E , can be readily determined. E is defined as

$$E = 1 - \frac{c_{eff}}{c_{in}} \quad (2)$$

where c_{in} and c_{eff} denote the influent and effluent concentrations, respectively. The extent of aerosol deposition within a filter is commonly described by the specific deposit, σ , defined as the volume of particles deposited per unit filter volume. The average σ , $\bar{\sigma}$, can be found to be

$$\bar{\sigma} = \frac{M/\rho}{V_B} p \quad (3)$$

where M is the total mass of aerosols collected, ρ_p the aerosol particle density and V_B , the filter volume.

The total volume of particles passing through the filter up to times t , is simply

$$M = \int_0^t u_s A c_{in} dt \quad (4)$$

where u_s and A are the superficial velocity and cross sectional areas of the filter.

The filtration results obtained are summarized in the following.

Unfluidized Beds. As an example, results obtained in runs F5 and F6, which were conducted at $0.75 u_{mf}$, are shown in Figs. 10 in the forms of E , Δp and $\bar{\sigma}$ vs. M . Run F5 was conducted without magnetic stabilization, while run F6 was operated at 100 Oersteds. The results of both runs demonstrate similar trends for both the magnetized and the unmagnetized beds as expected since applying magnetic field did not affect the unfluidized region.

The unfluidized beds provided high collection efficiencies (100 percent in both runs). Consequently, the pressure drop increased as aerosols were collected within the beds, that is, as the specific deposit increased with time, because of the decrease in bed voidage.

Unstably Fluidized Beds. The best example of an unstably fluidized bed is shown in Fig. 11. Runs F1 and F2 were conducted at $1.5 u_{mf}$ with no applied magnetic field. The low collection efficiencies resulted directly from the unstable bed conditions (i.e., bubbling).

Stably Fluidized Beds. Two experiments were conducted under stable conditions with an applied magnetic field of 150 Oersteds at gas velocities equal to

1.5 u_{mf} (run F7) and 2 u_{mf} (run F8). The two stably fluidized beds yielded high collection efficiencies (greater than 95 percent), as shown in Fig. 12a. The pressure drop across the beds, as shown in Fig. 12b remained constant at approximately the value of W/A (0.213 psi) even as the specific deposit within the beds increased with the increase of time. The large porosities in the stably fluidized beds (a result of bed expansion) enabled the collection of aerosols without significantly decreasing the void volume.

Defluidized Beds. Defluidized bed structures were obtained by increasing the air velocity in excess of the desired operating velocity. The flow rate was then reduced to the desired value.

Two types of defluidized beds were studied. The first bed structure (run F9 operated at 1.5 u_{mf} and a magnetic field of 100 Oersteds) was defluidized from a bubbling bed, while the second bed structure run F10 operated at 1.5 u_{mf} and 105 Oersteds) was defluidized from a stably fluidized bed. In carrying out run F9 the flow rate was initially increased to the transition velocity (approximately 2 u_{mf}), where the bed bubbled, and was then reduced to 1.5 u_{mf} , at which point the bubbling ceased. The bed structure in run F10 was obtained by increasing the flow rate to 2 u_{mf} (stably fluidized) and then decreasing it to 1.5 u_{mf} .

A comparison of the collection efficiencies from runs F9 and F10 is shown in Fig. 13a. Both runs exhibited high collection efficiencies after long hours of operation, although the unstable defluidized bed (run F9) gave comparatively lower collection efficiencies at the initial stages of filtration than the stable defluidized bed (run F10). The pressure drop of run F9, however, was found to be 25% lower than that of run F10 (see Fig. 13b). Both of these observations are consistent with the hypothesis that there

exists significant channeling effect in an unstable defluidized bed than a stable defluidized bed which, as shown before, behaves like a fixed bed.

A summary of the filtration characteristics of the unfluidized, the unstably fluidized, the stably fluidized, and the defluidized beds is given in Table 4.

Filtration Measurements at Marginal Stability. In the earlier work of Warrior and Tien (1986), it was found that under certain conditions, both collection efficiency and pressure drop of a magnetically stabilized fluidized filter declined significantly during the course of filtration when the extent of deposition became larger. The explanation for this decline was attributed to a reduction of inter-granule magnetic forces resulting from the aerosol saturation of the magnetic particles.

Warrior and Tien stated that their experiments in which significant decline in collection efficiency and pressure drop occurred were carried out in the stable fluidization regime. However, when their experimental calculations were compared with the reconstructed stability diagram of Fig. 9, it became obvious that the experiments were actually conducted in or near the marginally stable zone. It is conceivable that vulnerability to small disturbances of a fluidized bed operating in the marginally stable zone could very well result in the instability phenomena observed by Warrior and Tien.

Runs F11 and F12 were conducted at conditions similar to one made by Warrior and Tien (namely run 21 in their work). The collection efficiency and the pressure drop vs. time data are shown in Figs. 14a and 14b. The collection efficiency and the pressure drop both showed significant decline in the early stages of filtration. Visual observations revealed bed bubbling in conjunction with the decrease in efficiency and the pressure drop. In run

F11, the bed bubbled continuously after approximately 1 hr. Run F12, which showed bubbling at approximately 0.5 hr., ceased bubbling shortly thereafter. The bed granules in run F12 rearranged themselves in such a way that a metastable structure was obtained leading to higher collection efficiencies as the bubbles were eliminated. These results and observations suggest that bed structures in experiments which were operated near the zone of marginal stability exhibited unstable behavior. Both the collection efficiency and pressure drop declined and fluctuated with time. A reasonable explanation is the growth of instabilities caused by small disturbances in a metastable MSB structure (an MSB operating near the marginal state may be considered a metastable structure). The hypothesis of reduced inter-granular magnetic forces due to aerosol saturation advanced by Warrior and Tien is a valid argument but was not the cause of the instability theory observed.

Stability of MSB Structures with Aerosol Deposition. Four experiments (runs F13 through F16) were made to examine the effect of aerosol deposition on bed stability and filter performance. As stated earlier, Warrior and Tien suggested that aerosol saturation in stably fluidized beds may lead to the growth of instabilities due to reduced inter-granule magnetic forces.

The granular beds used in these experiments were presaturated with various quantities of aerosol (talcum powder) by mixing a desired amount of powder with the magnetite granules. It should be noted that, although this procedure led to a qualitative understanding of aerosol saturation and stability in the MSB structures, it did not allow a quantitative analysis because of the inability to accurately determine the amount of pre-deposited powder since a loss of powder inevitably occurred during the mixing process and during the initial stages of air flow through the bed.

Runs F13 through F16 were conducted in the stably fluidized regime at $2 u_{mf}$ with a magnetic field of 150 Oersteds. In run F13, before adding the magnetite to the system, 1000 mg of powder was mixed with the grains. Fig. 15a shows that the collection efficiency during run F13 remained extremely high as time increased. In addition, as shown in Fig. 15b, the pressure drop remained constant with time, and no instabilities were observed within the bed. Experiment F14 was conducted after mixing 3000 mg of powder with the magnetite grains. Figs. 16a and 16b show that a significant decline in both the collection efficiency and the pressure drop were observed after 0.8 cm^3 of particles had entered the bed at approximately 6 hrs into the experiment. Furthermore, in run F15, when 4000 mg of talc powder was mixed with the magnetite prior to the run, a significant drop in the collection efficiency and in the pressure drop occurred after 0.3 cm^3 of particles had entered the bed. This behavior corresponded to 2 hrs of operating time (see Figs. 17a and 17b). In runs F14 and F15, bubbling and spouting were observed at the same time as the collection efficiency and the pressure drop decreased.

The last experiment, run F16, was made to check the stability of an initially clean bed after long hours of operation. Fig. 18a shows that high collection efficiencies were obtained after 0.8 cm^3 of particles had entered the bed. In addition, the pressure drop data, shown in Fig. 18b, remained constant with increasing time. No bubbling was observed during the course of the experiment.

The results of experiments F13 through F16 suggest that aerosol saturation of the magnetite granules lead to the growth of instabilities. Bubbling was observed earlier in beds which had larger quantities of aerosol deposition.

Conclusions

In summary, a further study of magnetically stabilized fluidized beds (MSB) for their potential use in aerosol filtration was conducted. The following conclusions of the study are:

1. An MSB produced various bed structures (i.e., unfluidized, stably fluidized, unstably fluidized, and defluidized).
2. Defluidized beds were determined to be fixed beds because their linear pressure drop-flow rate data were represented by the Carman-Kozeny equation.
3. A stability diagram constructed from pressure drop flow rate experiments was similar to Rosensweig's (1979) with the addition of a marginal stability zone between the stably and unstably fluidized regimes.
4. Collection efficiencies in a stably fluidized bed were much higher than in an unstably fluidized bed and were comparable to collection efficiencies in unfluidized beds.
5. Stable defluidized beds, which were shown to be fixed beds, yielded high collection efficiencies while the pressure drop across the bed increased.
6. Because of low aerosol concentrations, inconclusive results were attained in the unstable defluidized beds.
7. Operating the MSB in the marginal zone of stability caused declines in both the collection efficiency and the pressure drop because of small disturbances leading to bed bubbling.
8. Large amounts of aerosol deposition within a stably fluidized bed led to bed bubbling as a result of inter-granule magnetic forces.

Nomenclature

A	cross sectional area of filter
c_{in}, c_{eff}	influent and effluent gas concentrations
E	total collection efficiency
H	filter height
M	total particles entered into the filter
u_{mf}	minimum fluidization velocity
u_s	superficial velocity
u_t	transition velocity
v_B	volume of filter
W	weight of filter grains

Greek Letters

ρ_p	aerosol particle density
σ	specific deposit
$\bar{\sigma}$	average specific deposit

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Table 1
Magnetite Properties

Approximate Molecular Formula	Fe_3O_4
Sieve Size	24-48 Tyler mesh
Number Average Diameter	570 μm
Density	4.60 g/cm^3
Minimum Fluidization Velocity	30 cm/s
Sphericity Factor	0.54

Table 2

Comparison of Experimental Pressure Drop Data in Magnetized Beds
with Pressure Drop Values Calculated from the Carmen-Kozeny Equation

Air Velocity u (cm/s)	Experimental Pressure Drop dP (psid)	Calculated Pressure Drop dP (psid)	% Error
Run P3			
0	0	0	-
15.4	0.034	0.033	2.9
23.1	0.049	0.049	-
30.8	0.065	0.066	1.5
38.5	0.081	0.079	2.5
46.2	0.092	0.095	3.3
53.9	0.099	0.108	9.1
61.6	0.105	0.123	17.1
69.3	0.111	0.138	24.3
77.0	0.117	0.154	31.6
Run P5			
0	0	0	-
15.4	0.072	0.066	8.3
23.1	0.106	0.094	11.3
27.0	0.124	0.110	11.3
30.8	0.141	0.126	10.6
34.7	0.158	0.142	10.0
38.5	0.174	0.157	9.8
46.2	0.204	0.179	12.3
Run P6			
0	0	0	-
15.4	0.068	0.063	7.4
23.1	0.099	0.090	9.1
27.0	0.116	0.105	10.4
30.8	0.133	0.119	10.5
34.7	0.149	0.134	10.1
38.5	0.164	0.142	13.4
46.2	0.186	0.170	8.6
50.1	0.204	0.185	9.3

Table 3
Filtration Experimental Runs with Stability
Diagram Location Numbers

Fig. 9 Location No.	Run No.	Magnetic Field H (Oersteds)	Air Velocity u/u_{mf} $u_{mf} = 30 \text{ cm/s}$
1	F1	0	1.5
1	F2	0	1.5
2	F3	100	1.5
2	F4	100	1.5
3	F5	0	0.75
4	F6	100	0.75
5	F7	150	1.5
6	F8	150	2.0
2	F9	100	1.5
5	F10	150	1.5
7	F11	100	2.0
7	F12	100	2.0
6	F13	150	2.0
6	F14	150	2.0
6	F15	150	2.0
6	F16	150	2.0

Table 4

A Comparison of Filtration Characteristics in MSB Structures

MSB Structure	Collection Efficiency	Pressure Drop History
unfluidized	very high	increases with deposition
unstably fluidized	fluctuates, decreases with time	constant
stably fluidized	very high	constant
stable defluidized	high	increases with deposition
unstable defluidized	*	increases with time

*results were not conclusive

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Fig 2

Fig 3

Case 1

Case 2

Fig 9

Fig. 10