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INTRODUCTION

This book contains four substantial articles covering a topical range of filtration and separation developments. The aims of this multi-author book remain the same as with previous volumes; to provide researchers and engineers with critical reviews and carefully considered opinions of recent works in the field of filtration and separation and adjoining areas. All the chapters in this edition bring their respective topics up to date, and examine the results from experiments techniques currently being used in the light of the most modern theoretical developments. The authors have provided definitive reports which utilise their own analyses of each subject, and they have exercised their own judgement as to which are significant papers and developments in their own aeras of expertise.

In the first chapter (Dust Filtration by Fabric Filters) Professor Leith and Dr Allen summarize the theories for pressure drop, particle penetration, and cleaning behaviour of fabric filters. It is clear from this that some similarities are beginning to emerge between the analysis of the formation of dust cakes and liquid cake filtration when macroscopic models for the two processes are compared. Of course, the microscopic features of each process can still be (and are) quite different. The authors then consider the use of information from the models to examine the scientific basis for filter design.

Flocculation is still becoming increasingly important for the successful operation of many types of solid/liquid separation equipment, and the effects of interactions between equipment geometry, hydrodynamic conditions and floc characteristics is little understood. However, a good fundamental understanding of flocculation is now emerging from the scientific literature and Dr Gregory brings much of this relevant work together in Chapter 2. In this chapter the forces between particles which determine colloidal stability, common methods of destabilizing suspensions, and flocculation kinetics (including effects of hydrodynamic interactions) are reviewed with particular emphasis on aqueous dispersions, although many of the concepts apply to non-aqueous systems.

In the third chapter Professor Fane discusses the latest developments and applications of Ultrafiltration, and shows quite definitively the importance of modelling the process as an aid to design. In this chapter the author gathers together a large reservoir of information and data pertaining to the nature of ultrafiltration membranes. Mechanistic aspects and models of the concentration polarisation phenomenon and for flux are highlighted, as is the fouling which is the demise of so many industrial ultrafiltration units.

The fourth chapter (Deliquoring by Expression - Theory and Practice) by Professors Shirato, Murase and Iwata detail the current thinking in the theoretical developments for expression. The text shows that these can be applied to industrial design problems, and introduces an analysis for continuous screw expression based on a batchwise variable-pressure, variable-rate expression theory.

The final chapter (Membrane Gas Separation) by Professors Sen Gupta and Sirkar offer a very comprehensive coverage of the modern literature, and their contribution focusses attention on the analysis of permeators with asymmetric membranes. Previous publications on gas permeator analyses have invariably concentrated on homogeneous membranes; even though commercial membranes are either asymmetric or composites theoretical analyses using models based on such membrane structures have appeared only recently. This chapter highlights some key features of the processes and their analyses.

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CONTENTS

DUST FILTRATION BY FABRIC FILTERS

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I. INTRODUCTION

The purpose of this chapter is to summarize theories for pressure drop, particle penetration, and cleaning behavior of fabric filters. This information is used to examine the scientific basis for filter design. In practice, however, the theories are not exact so that filter design must consider matters beyond the equations presented. No attempt is made to present an exhaustive review of the subject; rather, a picture is given of those ideas that workers from two research schools consider important, and that are likely to be developed further.

Fabric filters are often categorized by their cleaning method; however, in this chapter, classification is by the direction of gas flow and hence by the location of the dust deposit. The two possibilities are inside collection and outside collection.

Inside Collectors

Figure 1 shows an inside-collecting filter. During filtration, shown in Figure 1a, dusty gas passes upward into cylindrical or pocket shaped bags, that are closed at the top. Filtration occurs as dusty gas flows through the dust cake that builds on the inside bag surface. Cleaned gas passes out through the filter housing. Bags in a compartment such as that in figure 1 operate in parallel. Several compartments together comprise the fabric filter.

To clean the bags, the flow of dusty gas must be stopped. This is usually accomplished by closing poppet valves in the exit duct. Two cleaning techniques can be used with inside collectors: reverse flow and shaking.

In the first, shown in figure 1b, cleaned gas from the outlets of other compartments is forced backwards through the bags to be cleaned, collapsing them gently. The reverse gas flexes the bags and causes the dust deposit on the inside surfaces to crack, spall off, then fall to the hopper. Rings sewn around the circumference of the bags keep them open during cleaning so that gas can flow down the inside. After a minute or so the reverse gas stops; after an additional brief period for the dust to settle, the poppet valves reopen and filtration resumes.

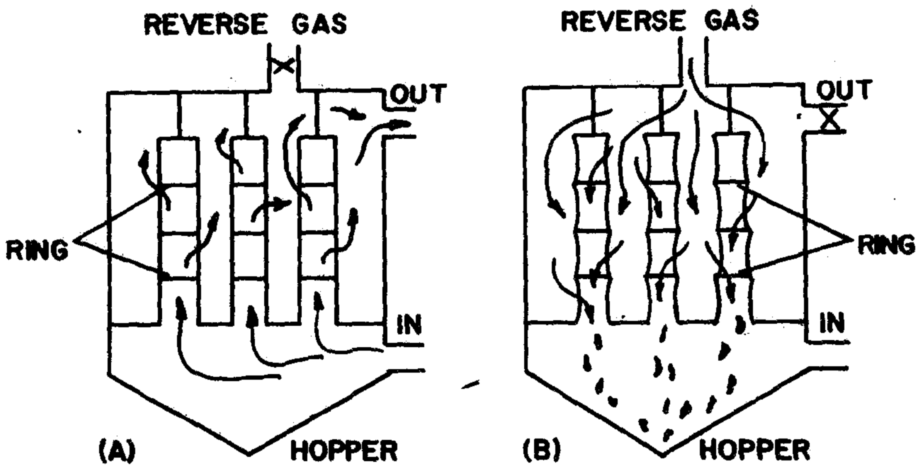

Fig. 1. Operation of an inside collecting filter: (a) during filtration, and (b) during cleaning.

In the second system the bags are cleaned by shaking. Bag tops are connected to an arm that oscillates with amplitude of 5-10 cm and frequency of 10-20 Hz. As with reverse gas cleaning, bag motion causes dust on the inside bag surfaces to separate from the fabric and fall to the hopper. Shaking is sometimes used together with reverse gas to remove dust more effectively than is possible with reverse gas alone.

Bags made from woven fabrics are generally used in these filters. Most large filters cleaned by reverse gas collect fly ash at electrical power stations. For this application, bags are made from glass fiber to withstand hot flue gas.

Outside Collectors

Figure 2 shows an outside-collecting filter. In the most common configuration, shown in figure 2a, dusty gas flows radially inward through cylindrical bags held open by metal cages placed within them. Dust collects on outside bag surfaces. Cleaned gas passes out the top of each bag to a plenum. Alternatively, pulse-cleaned bags can be flat and thin, arranged like vertical envelopes in a box. The open end of envelope bags, through which gas leaves the bags and through which pulse cleaning occurs, can be at the top or side.

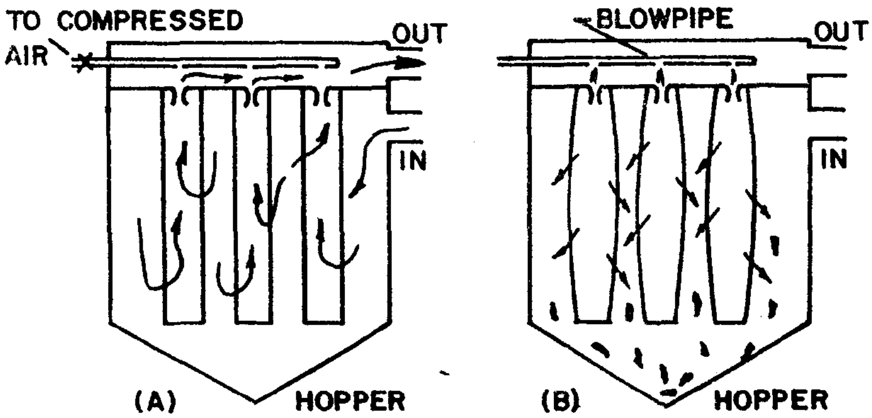

Fig. 2. Operation of an outside collecting filter: (a) during filtration, and (b) during cleaning.

Most outside collectors are pulse-jet filters, cleaned by injecting a momentary pulse of compressed air at the outlet of each bag as shown in figure 2b. This pulse snaps the bag open and drives collected dust away from the outside bag surface so it can fall to the hopper. The top of each bag may contain a venturi to entrain extra gas from the outlet plenum during cleaning. Pulse-jet cleaning takes a fraction of a second so it can occur on line, without interrupting gas flow to a compartment. If on-line cleaning is ineffective, bags can be pulse-cleaned off line by compartmenting the same way as for inside collectors.

Pulse-jet filters generally use felt fabrics rather than the woven fabrics used in reverse gas filters. The thick felt reduces dust penetration even if a dust cake is not fully developed, reducing penetration immediately after cleaning. Because the filtration velocity through a pulse-jet filter is several times higher than that through a reverse gas filter, a pulse-jet filter may be smaller and less expensive than a reverse gas filter. Pulse-jet collectors are the filters most commonly used to control industrial dust.

Cake vs. Media Filtration

A dust deposit builds away from the fabric as filtration proceeds. Dust collects close to the top surface of the cake, so that the particles collected most recently probably collect the dust that arrives soon thereafter. Until a continuous dust cake develops, filtration characteristics are affected by fabric properties as well as dust properties. "Cake filtration" occurs when the deposit becomes thick enough so that properties of the dust cake, rather than the supporting media, determine filtration characteristics. If the fabric-dust interface is smooth, the fabric has little effect on the dust deposit once a thin layer of dust has accumulated. If the fabric-dust interface is fuzzy or rough, considerable dust must collect before the deposit builds beyond the influence of the fabric and cake filtration begins. If the fabric surface is particularly fuzzy, a dust cake may not form before pressure drop is high enough to require fabric cleaning.

Woven fabrics have a smooth surface that requires little dust to form a cake. Furthermore, reverse gas cleaning is rather ineffective at removing this cake so that woven fabrics often retain considerable dust after cleaning. For these reasons, dust collection on woven fabrics is almost always on previously collected dust rather than on the fabric itself. Cake filtration characterizes these filters.

Cake filtration may or may not characterize a felt cleaned by pulse-jet. If the felt has a smooth surface as provided by a laminate or other surface treatment, if a woven fabric is used instead of a felt, or if the dust has high concentration and is coarse so a thick deposit develops between cleanings, then cake filtration may describe pulse-jet filter operation. Often, however, the felt surface is uneven and cleaning is frequent. In this case, a continuous dust cake may not form and fabric properties will affect filter performance. Confusion about pulse-jet filter performance often arises because they can be cake or non-cake filters depending on dust, fabric, and operation.

II. PRESSURE DROP

Pressure drop is the difference between static pressure upstream and downstream of the filter. It is the sum of the pressure drop across the filter housing and across the dust-laden fabric. Pressure drop across the housing is proportional to gas flow squared because flow through the housing is turbulent; pressure drop of this kind is described elsewhere (1). The discussion below concerns pressure drop across dust-laden fabrics.

Figure 3 shows an idealized plot of pressure drop, ΔP, against areal density of the dust deposit, W (mass of dust collected per unit fabric area). Before dust collects, pressure drop is due to the clean fabric alone. As dust collects, pressure drop increases--rapidly at first, then more slowly as a dust cake develops. After the cake has formed fully, pressure drop increases at a constant rate. When pressure drop becomes excessive the fabric is cleaned to remove dust, decreasing W. The time between successive cleanings of the same bag defines a filtration cycle.

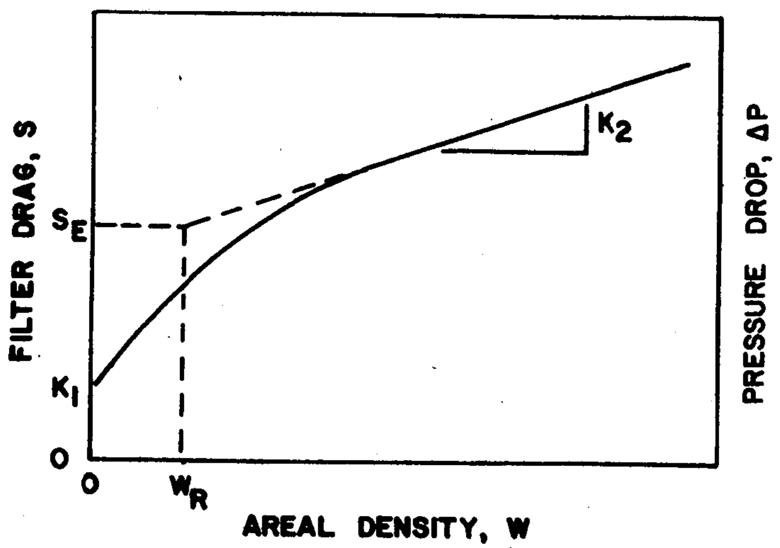

Fig. 3. Pressure drop and drag for a fabric filter vs. areal density of the dust deposit.

Pressure drop across the dust-laden fabric, ΔP, consists of that across the fabric, $K_1 V$, plus that across the dust, $K_2 V W$:

$$\Delta P = K_1 V + K_2 V W \quad . \tag{1}$$

Here, constant K_1 is the flow resistance of the clean fabric, constant K_2 is the specific resistance of the dust deposit first defined this way by Billings and Wilder (2), and W is the areal density of the dust deposit. Because flow is laminar through both the fabric and the dust deposit, pressure drop is proportional to filtration velocity. Division of equation (1) by V gives filter drag, S.

$$S = \Delta P/V = K_1 + K_2 W \tag{2}$$

A plot of drag, S, against areal density, W, is also given in figure 3. This equation is based on the assumption that the increase in drag with increasing areal density is linear; in fact, the increase is linear only after a dust cake has formed so that equation (2) gives drag that is too low.

This difficulty can be overcome by extrapolating drag determined experimentally to the residual areal density after cleaning, W_R, giving "effective" drag, S_E, as shown in figure 3. Equation (2) becomes

$$S = \Delta P/V = S_E + K_2 (W - W_R) \quad . \tag{3}$$

Equation (3) will give a higher value for drag than equation (2) because $S_E > K_1$; however, S_E cannot be calculated from theory and must be evaluated from experiments. This equation applies wherever S_E, K_2, W, and W_R are uniform.

Darcy's Law

Rudnick (3) has described the flow resistance of porous media. His work guides the discussion below.

Darcy's law characterizes flow through porous media (4),

$$V = B \, \Delta P/L \quad , \tag{4}$$

where V is superficial velocity, ΔP is pressure drop across the media, L is media thickness, and B is Darcy's permeability coefficient. Media thickness is difficult to measure but can be calculated from areal density if particle density, ρ_p, and porosity of the media, e, are known (5).

$$L = W/[\rho_p(1-e)] \quad (5)$$

Substitution of (5) into (4) yields, for pressure drop across the media,

$$\Delta P = K_2 \; V \; W \quad , \quad (6)$$

where $K_2 = 1/[B \; \rho_p \; (1-e)]$. The specific resistance of the dust deposit, K_2, is constant for a given media, dust, and porosity.

Evaluation of Specific Resistance, K_2

K_2 can be evaluated from theory or measured.

Stokes's Law Limit for K_2. For media comprised of isolated spheres, Stokes's law describes pressure drop. This relationship is realistic only if porosity approaches unity, e -> 1.

$$\Delta P = 3\pi \mu \;\; V \; D \; N/C' \quad , \quad (7)$$

where N is the number of spheres of diameter D per unit area in the media and C' is Cunningham slip correction factor. Substitution of (7) into (6), with $W = N \rho_p \pi D^3 /6$, yields

$$K_{2\ \mathrm{Stokes}} = 18 \; \mu/(\rho_p \; D^2 \; C') = 1/\tau \quad , \quad (8)$$

where τ is particle relaxation time (6,7).

Stokes's K_2, equation (8), defines pressure drop across an assembly of spheres uniformly distributed in space and far enough apart that they do not interfere with each other. It defines the lower limit for pressure drop across porous media. In actual media particles touch; flow around one affects flow around the next. This interaction causes pressure drop across media to be higher than predicted by Stokes's law. For tightly packed spheres and low porosity, pressure drop is much higher than predicted by equation (8).

To show the effect of porosity on pressure drop, Rudnick (3) defined a resistance factor, R, which gives actual K_2 when multiplied by Stokes's K_2.

$$K_2 = R \; K_{2\ \mathrm{Stokes}} = R/\tau \quad (9)$$

R is always greater than unity; it is large if porosity is low and approaches unity as porosity approaches unity.

Kozeny-Carman Equation. The Kozeny-Carman relationship (4), equation (10), is often used to describe pressure drop across a dust deposit. Flow is assumed to be through capillaries whose surface equals that of the particles comprising the media. Capillary volume is set equal to the media void volume. From these assumptions,

$$R = 2\, k_{CK}\, (1-e)/e^3 \quad , \qquad (10)$$

where the empirical constant $k_{CK} = 4.8$ for spheres and $k_{CK} = 5.0$ for irregular shapes.

As porosity approaches unity, equation (10) predicts that R approaches zero rather than unity as it should; therefore, if porosity is high, the Kozeny-Carman equation predicts R and pressure drop values that are too low. Billings and Wilder (2) cautioned that the Kozeny-Carman equation should not be used if $e > 0.7$. Carman (4) stated the equation should not be used if $e > 0.8$.

Rudnick-Happel Equation. Rudnick (3) used the Happel free-surface cell model (8) to determine resistance factor, R. In this model, each particle is assumed to be a sphere at the center of a cell. The volume of the cell is such that the porosity of each cell equals that of the media. Tangential stress within the fluid at surfaces where cells adjoin is set at zero. These assumptions allow exact solution of the Navier-Stokes equation, assuming the inertial terms are negligible. Resistance factor becomes

$$R = \frac{3 + 2(1-e)^{5/3}}{3 - 4.5(1-e)^{1/3} + 4.5(1-e)^{5/3} - 3(1-e)^2} \quad . \qquad (11)$$

Because it is derived from theory, equation (11) does not contain an empirical constant.

Figure 4 is a plot of resistance factor, R, against porosity, e, for the Kozeny-Carman and Rudnick-Happel equations. The figure shows agreement between the equations for porosities lower than 0.6, but for higher porosities the equations give divergent results. The Rudnick-Happel equation correctly predicts that R approaches the Stokes limit, unity, whereas the Kozeny-Carman equation incorrectly predicts that R approaches zero.

Rudnick (3) points out that equation (11) may or may not be applicable to non-spherical particles; furthermore, it may or may not be applicable to dust deposits comprised of small particles for which the surrounding fluid may not be considered continuous. With these reservations, equation (11) gives the best theoretical estimate of R.

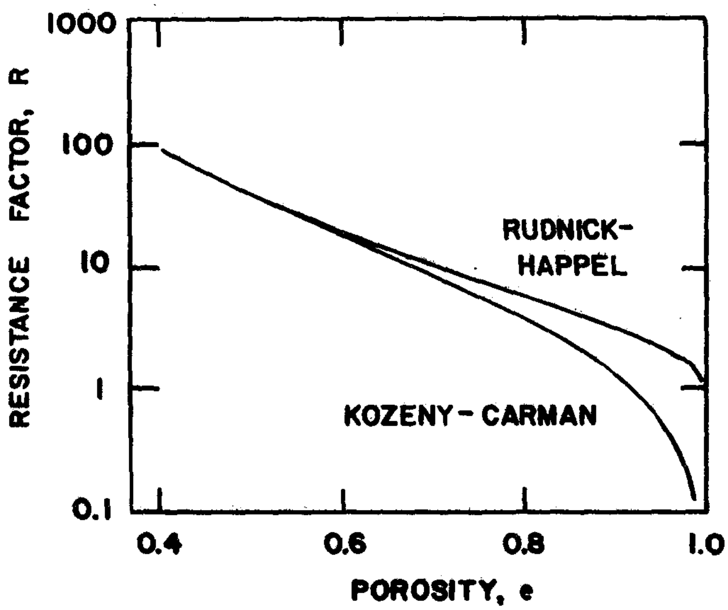

Fig. 4. Resistance factor vs. porosity for Rudnick-Happel and Kozeny-Carman models for resistance across porous media.

<u>Particle Size Distribution</u>. Equation (8) shows that K_2 is inversely proportional to particle diameter squared. If the particles in the dust cake are polydisperse, a mean particle diameter must be used. If the Kozeny-Carman relationship for R is used, equation (10), the correct diameter is the volume-surface mean, D_{vs}, also called the Sauter mean diameter.

$$D_{vs} = \sum n_i D_i^3 / \sum n_i D_i^2 \qquad (12)$$

Here, n_i is the number of particles with diameter D_i. If the Rudnick-Happel relationship for R is used, equation (11), Rudnick (3) has shown that the correct diameter is the volume-length mean, D_{vl}.

$$D_{vl} = [\sum n_i D_i^3 / \sum n_i D_i]^{1/2} \tag{13}$$

The difference between the volume-surface and volume-length means increases as particle size distribution becomes more polydisperse. As the geometric standard deviation of a log-normally distributed dust increases from 1 to 2 to 3, D_{vs}/D_{vl} increases (3) from 1 to 1.27 to 1.83.

Experimental Evaluation of K_2. The theory above shows K_2 depends strongly on media porosity. Unfortunately, porosity is difficult to measure, even in the laboratory. Billings and Wilder (2) suggest that dust deposit porosities range from about 0.4 to almost 1.0, corresponding to an 85-fold difference in pressure drop. Others report comparable porosity variations (9,10). These porosities were not measured, but were calculated from dust bulk densities or from measured pressure drops using the Kozeny-Carman equation with all terms known except porosity.

Rudnick (3) collected elutriated AC-fine dust on filter paper and woven fabric, then measured the resultant dust deposit porosities to be from 0.87 to 0.93. He found much better agreement between measured K_2 and K_2 predicted using the Rudnick-Happel model than between measured K_2 and K_2 predicted using the Kozeny-Carman model because measured porosities were above the range where the Kozeny-Carman equation is valid.

Because the porosity of a dust deposit is generally unknown and because R depends strongly on porosity, the theoretical equations for R given above cannot be used with confidence. An alternative to calculating R or K_2 from theory is to measure it. A K_2 value determined from measurements is called K_2' to distinguish it from a theoretical value. To find K_2', dust can be collected on a membrane filter and K_2' calculated from the increase in pressure drop, $(\Delta P_2 - \Delta P_1)$ with filter weight gain, $(M_2 - M_1)$.

$$K_2' = [A/V][(\Delta P_2 - \Delta P_1)/(M_2 - M_1)] \tag{14}$$

Here, A is surface area of the membrane filter.