1972

Granular filters for tertiary wastewater treatment

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72-26,920

HUANG, Jerry Yuan-Cheng, 1938-
GRANULAR FILTERS FOR TERTIARY WASTEWATER TREATMENT.

Iowa State University, Ph.D., 1972
Engineering, sanitary and municipal

University Microfilms, A XEROX Company, Ann Arbor, Michigan

Granular filters for tertiary

wastewater treatment

by

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A Dissertation Submitted to the

Graduate Faculty in Partial Fulfillment of

The Requirements for the Degree of

DOCTOR OF PHILOSOPHY

Department: Civil Engineering

Major: Sanitary Engineering

Approved:

Signature was redacted for privacy.

In Charge of Major Work

Signature was redacted for privacy.

For the Major Department

Signature was redacted for privacy.

For the Graduate College

Iowa State University
Of Science and Technology
Ames, Iowa

1972

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NOTATION

a	filter constant in Mintz's equation	-
A	number of pores susceptible to blocking	-
b	filter constant in Mintz's equation	-
c	filter coefficient constant in Ives' equation	L^{-1}
C	suspended solids concentration, mg/ℓ	ML^{-3}
C_c	critical filtrate quality, mg/ℓ of SS	ML^{-3}
C_o	influent suspended solids concentration, mg/ℓ	ML^{-3}
C_1	suspended solids concentration at layer l_1, mg/ℓ	ML^{-3}
C_2	suspended solids concentration at layer l_2, mg/ℓ	ML^{-3}
d	filter media size, mm	L
d_A	anthracite coal size, mm	L
d_e	equivalent media size, mm	L
d_s	sand size, mm	
E_{n-1}	steady-state at which exactly n-1 pores are clogged	-
E_n	steady-state at which exactly n pores are clogged	-
E_{n+1}	steady-state at which exactly n+1 pores are clogged	-
f	constant related to the distribution of particle size in the influent wastewater	-
g	gravity constant, ft./sec.^{2}	LT^{-2}

h	headloss, ft. water	L
H	total headloss through the filter bed, ft. water	L
H_c	critical (limiting) headloss through the filter bed, ft. water	L
H_o	initial headloss through the bed, ft. water	L
i	hydraulic gradient, dh/dl	-
js^2	dimensionless product relating to shape of grain and porosity (Camp's equation)	-
k	constant in Tchobanoglous and Eliassen equation	-
k_1, k_2	experimental constants in Tchobanoglous and Eliassen equation	-
K	headloss constant in Ives' equation	
l	depth of filter bed, or distance from surface of bed to section under study, in.	L
l_1	distance from surface of bed to layer 1, in.	L
l_2	distance from surface of bed to layer 2, in.	L
m	floc strength constant in Tchobanoglous and Eliassen equation	-
n	number of pores clogged	-
$N(t)$	number of pores blocked at the moment t	-
q	quantity of suspended solids deposited in the filter	ML^{-3}

q_u	ultimate quantity of solids that can be deposited	ML^{-3}
Q	wastewater flow rate, million gallons per day	L^3T^{-1}
r_o	initial removal rate in Tchobanoglous and Eliassen equation	T^{-1}
S	cross sectional area of filter bed	L^2
S_a	suspended solids accumulated inside filter pores	ML^{-3}
t	filter run length, or time from beginning of filter run, hr.	T
t_1	run length until filtrate quality declines to an unacceptable value, hr.	T
t_2	run length to the predetermined headloss, hr.	T
T	wastewater temperature, °C	-
v	flow rate, gpm/sq. ft.	LT^{-1}
v_o	initial interstitial velocity (t=0), gpm/sq. ft.	LT^{-1}
V_F	total volume of floc retained in the filter	L^3
x,y,z	experimental constants in Ives' equation	-
λ	impediment modulus, or filter resistance coefficient	L^{-1}
λ_o	initial impediment modulus (t=0), which is constant throughout the bed	L^{-1}
$\Theta(n)$	intensity function (or arrival rate) of the stochastic process	T^{-1}

σ	specific deposit, or the volume of particles deposited per unit volume of the filter bed, vol/vol	-
σ_u	ultimate specific deposit, vol/vol	-
α	constant in probabilistic model	-
β	constant in Ives' equation depending on the packing of the grains	-
γ	constant characterizing kinetics of tearing off of the particles in the probabilistic model	-
ε	porosity of the filter bed	-
ε_o	initial porosity of the filter bed	-
ϕ	filter coefficient constant in Ives' equation	L^{-1}
τ	kinetic viscosity of the fluid, sq. ft./hr.	L^2T^{-1}

ABBREVIATIONS

anthr	anthracite coal
BOD	biochemical oxygen demand
cm	centimeter
°C	degree(s) Celsius
COD	chemical oxygen demand
Cu. ft.	cubic feet
E.S.	effective size
ft.	feet
gal	gallon
gpd/sq. ft.	gallon(s) per day per square feet
gpm	gallon(s) per minute
gpm/sq. ft.	gallon(s) per minute per square foot
gr	gram
hr	hour(s)
in.	inch(es)
JTU	Jackson Turbidity Unit
lb	pound(s)
MG	million gallon(s)
MGD	million gallons per day
mg/ℓ	milligram(s) per liter
min	minute
ml	milliliter
mm	millimeter

NH_4-N	ammonia nitrogen
NO_2-N	nitrite nitrogen
NO_3-N	nitrate nitrogen
Org-N	organic nitrogen
Ortho-PO_4	orthophosphate
ppm	parts per million
psi	pound(s) per square inch
sec	second(s)
sq. ft.	square feet
SS	suspended solids
TOC	total organic carbon
total-PO_4	total phosphate
U.C.	uniformity coefficient
vs.	versus
μ	micron (= 10^{-6}cm)
%	percentage
$	dollar

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I. INTRODUCTION

With the exception of gravity sedimentation, granular filtration is the most widely used unit process for liquid-solids separation. Until recently its use was generally confined to the treatment of municipal and industrial water supplies. The primary reason for its recent adoption in tertiary wastewater treatment has been the need to upgrade effluents from conventional primary and secondary treatment plants. Such installations may use direct filtration of activated sludge or trickling filter final effluents, without the addition of chemical agents. Also, granular filters are employed in systems for phosphorous removal from secondary effluents and in physical-chemical systems for the treatment of raw and/or secondary wastewaters. In these latter cases chemical coagulation, flocculation and sedimentation precede the filters as in water treatment plants.

Filtration of secondary effluent is a difficult problem in many respects. If the secondary effluent contains a high suspended solids concentration, as many secondary effluents occasionally do, most of the material removed by a single graded sand filter is at or very near the surface of the bed, in which case the headloss increases very rapidly and most of the removal capacity of the filter bed is not utilized.

One approach to increasing the effective filter bed is the use of a dual media bed using a discrete layer of coarse anthracite coal above a layer of fine sand. This increases the efficiency of the filter as it provides for much greater utilization of the bed depth, using the fine sand only to remove the finer suspended solids. It is desirable to have the anthracite coal as coarse as possible to prevent surface clogging and the sand as fine as possible to promote high degrees of removals. However, the disparity in sizes cannot be too great lest overtopping of the coal by the sand result during backwashing. To ascertain the degree of mixing which will occur during the backwashing and its effects on subsequent filter performance, pilot plant studies must be conducted prior to filter design.

In essence, the design of a filter includes determination of: 1) type and size of filter media, 2) depth of filter media, 3) flow rate, 4) backwash rate, duration and timing of air and/or water backwash, air scour and surface wash, 5) type of chemical pretreatment and chemical dosage, and 6) expected run length. A design engineer is responsible for selecting these variables in order to yield a design which will produce an acceptable filtrate quality, preferably at least cost. The difficulty of arriving at an optimum design is due to the facts that these variables are interdependent and the present level of filtration theory can only semi-quantitatively relate the interdependence of these design variables.

All too often the design of a filter plant is performed according to "rule of thumb" experience, in which an over-design (uneconomical) or an underdesign (failure of system to meet quality or quantity objectives) usually results. Design information should be generated from pilot plant operation, from which a practical filter design can be reached.

This study can be introduced as a pilot filter plant study of the basic action occurring during wastewater filtration within a dual media filter bed and the effect of media type, size and depth of the filter bed and flow rate on the removal of pollutants. A method is to be developed and demonstrated for selecting the size and depth of anthracite and sand, taking into account the effect of inter-mixing. A rational method to design a filter plant will be presented and demonstrated by means of a case study of designing a tertiary wastewater treatment plant at Ames, Iowa.

II. OBJECTIVES AND SCOPE OF STUDY

The first objective of this study was to determine the effectiveness of granular filtration in reducing the pollution potential of the effluent from the Ames water pollution control plant, at which secondary treatment is provided by standard rate trickling filters. Particular attention has been devoted to the basic action of wastewater filtration within the filter bed and the effect of media size, depth of filter bed, and flow rate on the removal of turbidity and suspended solids. The change of headloss during the filter run, the floc storage at successive layers in the bed, and the effectiveness of an anthracite-sand dual media filter were also investigated.

The second objective of this study was to develop a method to select the size and depth of anthracite and sand as well as the flow rate and its corresponding run length based on the characteristics of the wastewater.

In order to accomplish these objectives, the study was conducted in several steps:

1. Operation of pilot single and dual media filters at the Ames water pollution control plant to determine the effectiveness of granular filtration in reducing the pollution potential of the final settled effluent.
2. Determination of the effects of using various media

sizes and their combinations on the filtrate quality and headloss development.

3. Determination of the effects of flow rate on the filtrate quality and headloss development.
4. Demonstration of how the size and depth of anthracite and sand are selected and demonstration of how the rational method can be applied in the filter plant design by a case study of designing a tertiary wastewater treatment plant at Ames, Iowa.

An extensive review was made of the literature to evaluate existing filter performance prediction models. It was found that wastewater filtration has been based mainly on experience gained in water works practice. Due to the variation of characteristics and SS concentration of the final effluent it is rather difficult, if not impossible, to predict the performance of wastewater filtration. Pilot plant operation prior to design cannot be overemphasized. In view of the many variations involved in filter design, the results of this study are not intended to be universally applicable. However, the methodology can be adapted for general practice.

III. LITERATURE REVIEW

The primary purpose of a literature review is to learn from the experience of others, to avoid making the mistakes which previous workers have committed, and to avoid repeating the works of earlier researchers. The knowledge accumulated from past works is like the foundation of a pyramid, while the current worker is like an artist adding pieces of stone onto that partially built pyramid. In order to make his own piece of work a valuable addition, the worker must have a thorough understanding of and familiarity with the previous works as well as with the creation of his new work. With this principle and purpose, the literature review will be conducted beginning with the physical nature of filtration, followed by mathematical models and ending with works about granular filters for wastewater filtration.

A. Physical Nature of Filtration

The analysis of forces that contribute to the transport and attachment and/or detachment of suspended particles in a deep-bed filter may help us to understand some of the reasons behind the performance of a filter under different circumstances. Several new theories have appeared since 1960, the most notable being the work of Ives (31-37). This section will summarize recent investigations into the mechanics and kinetics of deep-bed filtration.

It is generally agreed that filtration can be considered to comprise three principal mechanisms; transport, attachment, and detachment. Transport mechanisms move a particle into and through a filter pore so that it comes very close to a grain or existing deposits of particles; attachment mechanisms cause the particle to adhere to the surface of the grain or existing deposits; and detachment mechanisms occur due to the action of hydrodynamic forces of the flow such that a certain part of the previously adhered particles less strongly linked to the others is detached from grains or previous deposits.

1. Transport mechanisms

The fluid flow patterns in a filter bed of randomly packed grains are too complex to analyze in a precise geometric way. However, in the grain sizes of interest in rapid granular filtration (0.4 to 2 mm), at the flow rates of interest (2 to 8 gpm/sq. ft.), with water at a temperature between 0°C and 30°C (viscosity 0.018 to 0.008 poises) no evidence of departure from laminar flow conditions has been found. That is, there has been no separation of the flow boundary from the grain surfaces, i.e., no vortex formation, and the relation between headloss and flow rate has remained linear. This means that Poiseuille flow has dominated and the fluid inertia terms of the Navier-Stokes equations have been negligible. Therefore, fluid velocity is zero at the boundaries, i.e., grain surfaces, and is maximum at the center of a pore. Consequently transport mechanisms have to provide forces to move particles out of their flow stream lines into the proximity of the grain surfaces where fluid flow velocities are small, tending toward zero at the boundary.

Important transport actions are screening, interception, inertial forces, sedimentation, diffusion and hydrodynamic forces. The suspended solids removal efficiency and the type of transport actions which dominate depend on the sizes of the suspended solids and their distribution in any particular water or wastewater filtration. Yao, et al. (96) find that there exists a size of suspended solids for which the removal efficiency is minimum. This critical suspended solids size is about 1μ . For suspended solids larger than 1μ , removal efficiency increases rapidly with particle size. Removal is enhanced by transport forces of sedimentation and/or interception. For suspended solids smaller than 1μ , removal efficiency increases with decreasing particle size. Removal is made possible by diffusion. It is useful to note that many suspended solids of interest in water and wastewater treatment are about 1μ in size or smaller.

Ison and Ives (30), O'Melia and Stumm (68), and Selmeczi (75) have prepared excellent reviews concerning these mechanisms.

2. Attachment mechanisms

Attachment of particles to grain surfaces or existing deposits of particles has been generally attributed to physico-chemical and molecular forces. In many cases particles are attached by molecular bridges, such as synthetic, natural or hydroxide polymers, which are specifically absorbed to suitable sites - at one end on the particle, at the other end on the grain surface. Sometimes specific ions act as chemical links, such as calcium linking kaolinite and polyacrilamide. Yao, et al. (96) observe that attachment can be improved by using the optimum polymer dose determined in the coagulation jar tests. These chemicals probably enhance attachment in filtration by absorption to produce charge neutralization and/or bridging (68, 96). It is suggested that when conventional filters fail to produce efficient filtration, effective improvements can be made by altering the chemistry of the system - applying the optimum coagulant dose.

Major attachment actions as suggested by various researchers are friction, gravity, electrokinetic interactions, molecular forces and surface tension. Detailed discussions of these various forces have been presented by Ives and Gregory (36), O'Melia and Stumm (68), and Selmeczi (75).

3. Detachment mechanisms

This is a controversial subject among research workers in the area of filtration. One group of research workers, primarily Mintz (58-61), considers that deposits accumulated in the depth of a filter medium have an unequally strong structure. Under the action of hydrodynamic forces due to the flow of water through the medium, which increase with increasing headloss, this structure is partially destroyed. A certain part of previously adhered particles less strongly linked to the others is detached from the grains. Consequently, as the deposits accumulate they become unstable and parts of them are torn away by the flow, to go back into suspension in the pores.

There are two supporting pieces of evidence for this. First, polymers added to an in-flowing suspension cause a clearer filtrate to emerge (35). This is evidence that additional strength is given to the deposit structure by polymer binding. Second, visual observations by Mintz, et al. (61) on a shallow model filter in the absence of polymers show particles emerging in the filtrate which are larger than those in the suspension entering the filter. This is cited as evidence for the detachment of aggregates from the deposits in the pores.

The study of rod filters by Stein (81) showed that previously deposited floc of weaker strength was torn off as the interstitial velocity and shearing forces increased due to a decrease in the pore space. Weak particles in the sheaths were torn off and replaced by stronger particles that adhered more firmly. As a consequence, the surface of a thick sheath became stronger than the underlying floc. However, as the shear was further increased while the channels became smaller, the floc mass on a few of the rods sloughed off and was intercepted by the constrictions below. Stein observed this breakdown in the rod filter on many occasions and concluded that it probably also takes place in a granular filter bed.

The detachment phenomenon was also observed by Tuepker and Buescher (92) when flow rate was suddenly increased.

Another group of researchers, Ives (35), Lerk (46) and Mackrle and Mackrle (56) opposed this detachment mechanism. They considered that, as the interstitial velocity increases, and as the surface available in the filter pores and the amount of divergence and convergence of flow diminish due to the deposits accumulating in the pores, there is a reduction in the probability of particles being brought to a surface for adherence. Ives (35) quoted Stanley's (80) observations that even in the presence of a continuously flowing suspension, radioactively-labelled iron floc was not detached from its original place of deposition in the filter.

Camp (7) noted the rods in Stein's filter were not in contact with each other, whereas the grains in a granular filter bed rest on the grains below. There is much less freedom for breakdown in a granular bed, therefore, than in Stein's rod filter. Camp disagrees with Stein's conclusion, reasoning that such a breakdown, if significant in a granular bed, should be reflected in the shape of the headloss curves. The hydraulic gradient should decrease in the filter at the depth where the breakdown occurs and increase at the depth where the floc mass is again intercepted. Camp found nothing to indicate significant breakdown - in almost all cases, the rate of increase in headloss at all depths is greatest at the end of filter runs.

Evidences presented and interpretations made by both sides to support their argument can be found in the works of Mintz (59) and Ives (35).

The disagreement between the two groups of research workers concerning the role of detachment is not yet resolved.

The ultimate goal of studying the physical nature of filtration is to formulate a mathematical model which can be used to describe the time-space variation and build-up of material within the filter. In the next section, the development of mathematical models will be reviewed.

B. Mathematical Models

Development of mathematical models concerning particle deposition inside the pores of a filter media has been studied quantitatively by two separate groups of workers with little communication between them. One group is represented by chemical engineers concerned with the recovery of chemical solids and their clogging of filter cloths, of which comprehensive studies have been conducted by Ruth (74), Heertjes (23), and Tiller (87-89). Their results are expressed by ordinary differential equations with respect to time, and they have assumed that the effect of filter cloth thickness was negligible because it was thin. Since this thickness can often be equivalent to several hundred particle or pore diameters, this assumption might be re-examined in the light of the results from the other group of workers, mainly civil engineers, who investigated the clarification of water by sand filters (31-38, 56, 58-61, 69, 81) and who developed partial differential equations with respect to time and filter depth.

All mathematical models concerning filtration can be divided into two parts: one relating to rate of clarification or the theory of suspension removal, the other relating to rise in headloss due to filter clogging.

1. Rate of clarification

Rather voluminous works which concern the theory of suspension removal have appeared in the literature. One group of research workers based their equations on the rather idealized assumptions which govern the removal behavior which occurs within the filter pores. Another group of workers consider the attachment-detachment phenomena in the course of filtration as a stochastic process, in which operations research techniques such as queueing theory and Markov chains are applied in developing their models. Still another group of workers have been interested in an empirical approach, in which extensive experimental data are collected to develop empirical equations relating the media depth and size and gradation, flow rate and water characteristics to the filter effluent quality. In order to have a better understanding of this development, this review will proceed by grouping the discussions according to the approaches adopted by various research workers for describing the rate of clarification in a filter run.

a. Idealized mathematical model If the volume of the floc particles does not change significantly as they pass from above the filter into the bed and are deposited on the grains to remain throughout the run, it is evident that the volume removed from the water is equal to the volume deposited in the bed. This relation was first stated by

Iwasaki (38) in 1937 as follows:

$$\frac{\partial \sigma}{\partial t} = -v \frac{\partial C}{\partial \ell} \qquad (1)$$

where σ is the specific deposit, i.e., the volume of particles deposited per unit volume of filter bed, t is the time from the beginning of the filter run, C is the suspension concentration, v is the flow rate, and ℓ is the distance from the top of the bed to the section under study. In Eq. 1, $\partial\sigma/\partial t$ denotes the time rate of change of the deposit ratio at a particular depth, ℓ, and a particular time, t; $\partial C/\partial \ell$ symbolizes the corresponding rate of decrease in volumetric concentration of floc in the water. This equation is valid regardless of the manner of deposit within the bed, but its validity is restricted to cases in which the volume of floc already deposited is not reduced by loss of water. Subject to this restriction, Eq. 1 may be used to compute $\partial C/\partial \ell$ at a particular depth and time, provided that σ can be determined as a function of time during a run. Eq. 1, called the mass conservation equation, is regarded as one of two fundamental equations from which mathematical models of the rate of clarification have been developed.

Iwasaki (38) also proposed a relation between the rate of removal of floc and the concentration, as follows:

$$\frac{\partial C}{\partial \ell} = -\lambda C \qquad (2)$$

in which the proportionality factor, λ, is called the impediment modulus. This is the second fundamental equation from which mathematical models of rate of clarification have been developed.

Iwasaki assumed that λ increased linearly with σ because of the increased surface area available for adhesion. Stein (81) realized that a careful experimental study made a few years previously by Eliassen (15) keyed in with Iwasaki's mathematics. Stein modified the equation so that λ increases linearly, then decreases nonlinearly with the amount of clogging.

In 1951, Mintz (58) proposed a new theory based on the physical consideration that a change of particle concentration at each individual layer of the bed is the overall result of two opposing processes: extraction of particles from the water and their attachment to filter media under the action of adhesive forces, and detachment of previously adhering particles under the action of the hydrodynamic energy of the stream. Mathematically, these are expressed by a system of differential equations:

$$-\frac{\partial C}{\partial \ell} = bC - \frac{a\sigma}{v} \qquad (3)$$

$$-v\frac{\partial C}{\partial \ell} = \frac{\partial \sigma}{\partial t} \qquad (4) = (1)$$

where a and b are filtration constants depending on the flow rate, filter media, physico-chemical properties of the suspension, and other factors determining the filtration conditions.

It is evident that one of Mintz's equations, Eq. 4, is identical to Iwasaki's concept, Eq. 1. In fact, Mintz's ideas were similar to Iwasaki's, but he saw the filtration process as being composed of a constant deposition in the filter pore together with a shearing away of existing deposits. Mintz drew on Eliassen's (14) experimental data to support his theory, but he was apparently unaware of Iwasaki's or Stein's work.

In 1955, Ornatskii, et al. (69) reported an extensive study on the clogging of sand beds by clay suspensions. Ornatskii, et al. assumed that λ decreases linearly with σ, being inversely proportional to interstitial velocity. Their final equations are for C and σ.

$$\frac{C_{(\ell,t)}}{C_o} = \frac{\exp\ (\text{constant} \cdot v_o\ C_o\ t)}{\exp\ (\lambda_o \ell) + \exp\ (\text{constant} \cdot v_o\ C_o\ t) - 1} \tag{5}$$

$$\sigma_{(\ell,t)} = \text{constant} \cdot \lambda_o \cdot \frac{1-\exp\ (\text{constant} \cdot v_o\ C_o\ t)}{1-\exp\ (\text{constant} \cdot v_o\ C_o\ t) - \exp\ (\lambda_o \ell)} \tag{6}$$

where C_o is the influent suspended solids concentration, λ_o is the initial impediment modulus (t = o) which is constant throughout the bed, and v_o is the initial interstitial velocity (t = o). Ornatskii's propositions led to a fairly straightforward statement of how the filtrate quality, C, and clogging, σ, depend on both filter depth, ℓ, and time of filtration, t.

The brothers Mackrle (56) presented their hypothesis dealing with physico-chemical forces between filter grains and suspended particles. They attributed the adhesion in the filter to Van der Waals forces, and so quantum mechanics entered the picture. However, upon examination, their final correlation turns out to be more simply described as an inverse proportionality between λ_o and v_o, thus confirming Ornatskii's main assumption.

Camp (7), in 1964, presented a balanced account of theory, experiment, and the application of his own, Stein's (81), and Eliassen's (14-15) work on rapid filtration. He proposed an equation relating the hydraulic gradient, i, i.e., dh/dℓ, the headloss, dh, per unit of length, dℓ, in the direction of flow, to the specific deposit, σ, as follows:

$$\frac{id^2}{\frac{js^2\tau}{g}v} = \frac{(1-\varepsilon_o+\sigma)^2}{(\varepsilon_o-\sigma)^3} \frac{1}{\left[\sqrt{\frac{\sigma}{3(1-\varepsilon_o)}+\frac{1}{4}}+\frac{\sigma}{3(1-\varepsilon_o)}+\frac{1}{2}\right]} \qquad (7)$$

where d is media size, js^2 is the dimensionless product relating to shape of grain and porosity, g is the gravity constant, and τ is the kinematic viscosity of the water. This equation is based on the assumption that the removal is accomplished by means of a floc sheath on the grain.

Eq. 7 implies that if the hydraulic gradient (i), the physical characteristics of the filter bed (d, js^2, ε_o), flow rate and kinetic viscosity (τ) of the water are known at any time during the filter run, the specific deposit may be computed by means of Eq. 7.

In the early 1960s, Ives (33) reconciled the two conflicting assumptions of Stein and Ornatskii by suggesting that λ first increases with σ because of the increased surface area, then decreases because of the increased interstitial velocity and the smoothing of flow paths. His theoretical equation for λ is:

$$\lambda_{(\ell,t)} = \lambda_o + C\sigma_{(\ell,t)} - \phi \frac{\sigma^2_{(\ell,t)}}{\varepsilon_o - \sigma_{(\ell,t)}} \qquad (8)$$

where λ_o is the initial impediment modulus, ε_o is the initial porosity of the bed, i.e. the porosity of the empty bed, and C and ϕ are constants for all values of ℓ and t. It should be noted that this equation was developed based on the filtration of uniform sized particles on uniform sized filter media.

In 1967, Ives (34) presented a modified form of the equation of continuity:

$$- \frac{\partial C}{\partial \ell} = \frac{1-\varepsilon}{v} \frac{\partial \sigma}{\partial t} \tag{9}$$

where ε is the porosity of the deposited particles.

Ives assumed that for a given suspension at a given temperature, the filter efficiency was dependent on the surface area available for particle deposition and on the flow rate past such surfaces. From considerations of the geometrics of a coated sphere and a coated cylinder, and the velocities through a clogging pore, he proposed the equation:

$$\frac{\lambda}{\lambda_o} = (1 + \frac{\beta}{\varepsilon})^y \ (1 - \frac{\sigma}{\varepsilon})^z \ (1 - \frac{\sigma}{\sigma_u})^x \tag{10}$$

where β is a constant dependent on the packing of the grains, σ_u is the ultimate specific deposit, and x, y, z are empirical indices.

A study on multilayer filtration by Mohanka (62) showed that as σ approaches σ_u or as velocity increases in the pores, the exponent of the third term, x, is to be raised to a power (x>1) to satisfy the λ versus σ experimental curves. All the parameters, λ_o, β, σ_u, x, y, z, in Eq. 10 were evaluated from the experiment.

In 1970, Tchobanoglous (84) and Tchobanoglous and

Eliassen (85) conducted the first intensive study of filtration of secondary final effluent. They found that screening proved to be the principal removal mechanism operative in the filtration of settled sewage effluent and proposed a modified first order equation describing the rate of change of suspended solids concentration with filter depth:

$$\frac{dC}{d\ell} = \left[\frac{1}{(1+k\ell)^{f}}\right] r_0 C \left(1 - \frac{q}{q_u}\right)^{m} \tag{11}$$

where r_0 is the initial removal rate per inch depth, q is the quantity of suspended solids deposited in the filter, q_u, is the ultimate value of q, and k, f and m are constants. Constant m is related to floc strength. Initially, when the amount of material removed by the filter is low, $q \simeq 0$, $(1 - q/q_u)^m \simeq 1$, and Eq. 11 becomes

$$\frac{dC}{d\ell} = - \left[\frac{1}{(1+k\ell)^{f}}\right] r_0 C \tag{12}$$

When f is equal to zero, the term within the brackets is equal to one; under this condition Eq. 12 represents a a logarithmic removal curve, which is identical to Iwasaki's equation, Eq. 2. When f equals one, the value of the term within the brackets drops off rapidly in the first few inches and then more gradually as a function of filter depth. Therefore, it appears that the exponent f may be related to the distribution of particle size in the influent.

As the upper layers begin to clog, the term $(1 - q/q_u)^m$ in Eq. 11 becomes zero and the rate of change in concentration with depth is equal to zero. Thus, at the lower depths the amount of material removed is essentially zero.

Tchobanoglous and Eliassen (85) presented a graphical method to determine the initial removal rate, r_o, and the constant k, f and m. They found that $r_o = 1.0$ per in.; $k = 2.0$; $f = 2.0$ and $m = 2.0$.

b. <u>Probabilistic mathematical model</u> Instead of advancing the clarification equation (rate of removal per depth or per time) as research workers in previous sections did, a different group of workers introduced what is called "kinetics equation" - kinetics of clogging in the filter pores.

In 1961, Bodziony and Litwiniszyn (4) touched off the study of the process of clogging in the filter pores using a probabilistic model approach. Litwiniszyn (50) considered the process of clogging in the filter as a certain stochastic process. He proposed that the kinetics of the process of clogging were described by the equation:

$$\frac{\partial N(t)}{\partial t} = \alpha C[A - N(t)] \qquad (13)$$

where $N(t)$ is the concentration of the pores blocked at the moment t by suspended solids, α stands for a certain constant, C denotes the concentration of suspended solids, and A is the number of pores in the volume unit susceptible to blocking. Eq. 13 states that the rate of clogging, $\partial Q(t)/\partial t$, is proportional to the momentary concentration of free pores still susceptible to blocking, $[A - N(t)]$, and solid concentration, C. This equation can be considered a law resulting from averaging the distribution which determines the stochastic process.

Assuming the process of clogging to be a process of "pure births", he assumed that after a time interval the state E_n, in which there are n pores clogged, may transform only into the state E_{n+1}, in which there are n+1 pores clogged, then the probability of arriving at the state E_{n+1} after the time interval Δt equals $\theta(n)\Delta t$. $\theta(n)$ designates the so-called intensity function (or mean arrival rate) of the process, which is assumed to be dependent on n and independent of time, t. Mathematically,

$$\theta(n) = \alpha(A-n) \qquad (14)$$

where n is the number of pores clogged which runs the values of integers from zero to A.

The intensity function, as defined by Eq. 14, means that the probability of suspended particles being seized by the pores of the filter media is less as the number of non-blocked pores decreases. When all pores - the number of which in a volume unit of the medium equals A - are blocked, such probability falls to zero.

The fundamental system of equations of the process of "pure births" is of the form

$$\frac{d\,P_o(t)}{dt} = -\,\Theta(o)\;P_o(t)$$

$$\frac{d\,P_n(t)}{dt} = -\,\Theta(n)\;P_n(t) + \Theta(n-1)\;P_{n-1}(t) \tag{15}$$

$$\text{for } n = 1, 2 \cdots\cdots A$$

where $P_n(t)$ is the probability that exactly n pores are clogged at the moment t.

Introducing the intensity function, Eq. 14, into the system of Eq. 15 yields:

$$\frac{d\,P_o(t)}{dt} = -\,A\alpha P_o(t)$$

(16)

$$\frac{d\,P_n(t)}{dt} = -\,\alpha(A-n)\;P_n(t) + \alpha[A - (n-1)]\;P_{n-1}(t)$$

$$\text{for } n = 1,2 \cdots\cdots A$$

At the initial moment, t = 0, at the beginning of the filter run, the following initial condition is satisfied:

$$P_o(o) = 1$$

(17)

$$P_n(o) = 0 \qquad \text{for } n = 1,2 \cdots\cdots A$$

It can now be determined - from the system of Eq. 16 and the initial condition, Eq. 17 - that

$$P_o(t) = e^{-A\alpha t}$$

$$P_1(t) = Ae^{-A\alpha t+\alpha t}\;[1-e^{-\alpha t}] \qquad (18)$$

$$P_2(t) = \frac{1}{2}\,A(A-1)\;e^{-A\alpha t+2\alpha t}\;[1-e^{-\alpha t}]^2$$

In a later advance, Litwiniszyn (49) took into consideration the phenomenon of swelling of clogging particles, from which he developed his modified probabilistic model. In the modified model due attention is paid to the fact that the velocity of fluid flow carrying the particles varies with the changes in the dimensions of the section the fluid flows through. The changes in the section are caused by the clogging particles in the pores. There are

$$\frac{N(\ell,t)}{\partial t} = \alpha[A-N(\ell,t)]\ C(\ell,t) + G[N(\ell,t),t] \tag{19}$$

in which $G[N(\ell,t),t]$ describes the rate of concentration increase, $\partial N(\ell,t)/\partial t$, due to the swelling of the deposited particles.

In 1966, the concept of a detachment mechanism was incorporated into the work of Bodziony and Kraj (3). They postulated the equation of kinetics of the attachment-detachment process to be of the following form:

$$\frac{\partial N(\ell,t)}{\partial t} = \alpha[A - N(\ell,t)]\ C(\ell,t) - \gamma N(\ell,t) \tag{20}$$

where $\gamma > o$ represents a constant characterizing the kinetics of tearing off of the particles. Eq. 20, conjointly with the equation of mass balance in the form

$$\varepsilon\frac{\partial C(\ell,t)}{\partial t} + \varepsilon V\frac{\partial C(\ell,t)}{\partial \ell} = -\frac{\partial N(\ell,t)}{\partial t} \tag{21}$$

constitutes a set of equations with two unknowns: $C(1,t)$ and $N(1,t)$. ε in Eq. 21 denotes the porosity and v is the velocity of flow. It is not an easy task to find the solution of this set of equations. Thus, its usefulness is limited.

Along the same attachment and detachment argument, Litwiniszyn (51) advanced his model by incorporating the detachment mechanism while he was developing his stochastic model. There are two possibilities: the possibility of the system passing from the state E_n to the state E_{n+1} (attachment occurs) and the possibility of the system passing from the state E_n to the state E_{n-1} (detachment occurs). It is assumed that both processes, i.e. $E_n \to E_{n+1}$ and $E_n \to E_{n-1}$ are of a random nature. It is reasoned that within the time interval Δt, the probability of transition $E_n \to E_{n+1}$ is proportional to the number of free pores susceptible to filling in the volume unit of filter media, i.e. equal to $\alpha(A-n)\Delta t$; the probability of transition $E_n \to E_{n-1}$ is proportional to the number of particles trapped in a volume unit of the filter media, i.e. equals $\gamma n\Delta t$.

It may happen that within the time interval Δt, the state of the system will not be changed. The probability of such an event may be expressed by the formula

$$[1 - \alpha(A-n)\Delta t - \gamma n\Delta t]$$

Denoting by $P_n = P_n(t)$ the probability that at the moment t the system is in the state E_n, and taking into consideration the assumptions formulated above results in the following equation:

$$P_n(t+\Delta t) = \alpha[A-(n-1)]\Delta t\, P_{n-1}(t) + [1-\alpha(A-n)\Delta t-\gamma n\Delta t]\, P_n(t) + \gamma(n+1)\Delta t\, P_{n+1}(t) \tag{22}$$

Since A is the maximum number of particles which may be trapped in a volume unit of the filter media,

$$P_n(t) = 0 \qquad \text{for } n > A$$

As $\Delta t \to 0$, Eq. 22 becomes

$$\frac{dP_n(t)}{dt} = \alpha[A-(n-1)]P_{n-1}(t) - [\alpha(A-n)+\gamma n]\, P_n(t) + \gamma(n+1)\, P_{n+1}(t) \tag{23}$$

$$\text{for } n = 1,2 \cdots\cdots A$$

Since the system can pass from the state E_o only to the state E_1, it has in particular

$$\frac{d\,P_o(t)}{dt} = -\alpha A\,P_o(t) + \gamma P(t) \qquad (24)$$

The system of differential Eqs. 23 and 24 describe the linear stochastic process: births-deaths, the expression $P_n = P_n(t)$ defining the probability that at the moment t there are n particles trapped in a volume unit of the filter media. This linear model may be considered to be the first approximation of the kinetics of the attachment-detachment process.

Along the same concept and approach, further studies have been advanced by Litwiniszyn (48, 52-53) and Kraj (42-43). Unfortunately, little experimental data has been presented by these workers to test the validity of their highly sophisticated equations. Besides, general solutions for these systems of differential equations are rather difficult, if not impossible. The difficulty in obtaining the experimental constants, i.e. α and γ, reduced its appeal as a popular tool.

c. Empirical mathematical model Acknowledging the fact that complex and unpredictable affects on filter performance are caused by various characteristics of influent suspended solids in addition to the filter physical characteristics and operating conditions, several investigators (5, 9, 22, 26, 28-29, 76) have tried to find a way to correlate measurable properties of a water suspension with the basic design criteria to be used for sand filters so that both effluent and economical facilities for the treatment of water can be developed. One approach to developing such a correlation has been to determine the filtrability of a water source. Filtrability has been defined as the ease with which a water can be passed through a given filter and the effectiveness with which the solids are removed in the filter.

Considerable work has been done on the use of membrane filtration tests for defining quantitatively the filter-clogging properties of natural waters. The work of Boucher (5) is representative of this group. However, the membrane filtration test fails to indicate the extent of depth removal, which is encountered in granular filtration. Works of Hudson (28-29), Gamet and Rademacher (22), Hsiung and Cleasby (26), and Cleasby (9) present some ways to predict the performance of a granular filter.

Conley and Hsiung (11) extend the empirical model to the design and application of multimedia filters. They propose that the variables that affect filtration efficiency and headloss, such as flow rate, media size, filter depth, and amount of suspension in the influent can be arranged in grouped terms. Experimental data are then used to establish the proper exponentials to be used with these grouped terms.

However, agreement among the various investigators and the practical application of an empirical model is yet to be reached.

2. Headloss

If filter media are clarifying suspensions as they flow through, it follows that the pores of the media accumulate deposits which cause a loss of permeability or an increased flow resistance. The approach used most commonly to determine the headloss in a clogged filter has been to compute it with a modified form of the equations used to evaluate the clean water headloss. The formulations most often used are those proposed by Kozeny (41), Fair and Hatch (20), and Rose (73). In all cases, the difficulty encountered in using these equations is that the porosity must be estimated for various degrees of clogging. The complexity of this approach, unfortunately, makes most of these equations of little use.

According to Ives (33), the hydraulic gradient through a filter bed is related to the volume of the deposited flocs by

$$\frac{dH}{d\ell} = \frac{dH_o}{d\ell} + K\sigma \tag{25}$$

where H_o and H are the total headloss before and after floc accumulation, respectively, and K is a headloss constant which is assumed to be independent of depth. Eq. 25 can be integrated to give the total headloss across the filter; thus,

$$H = H_o + K\int_o^{\ell} \sigma d\ell \qquad (26)$$

The initial headloss, H_o, can be computed using one of the aforementioned equations; and the second term on the right-hand side represents the integrated headloss across the whole filter bed depth, l.

This relationship was developed further in a report by Engineering-Science Inc. (16). The total volume of floc retained by the filter is given by

$$V_F = S\int_o^{\ell} \sigma d\ell \qquad (27)$$

where S is the cross sectional area of the filter bed.

Substituting Eq. 27 into Eq. 26 yields

$$H = H_o + \frac{K}{S} V_F \qquad (28)$$

Therefore, for a specific media and flow rate, the total headloss depends only on the volume of floc retained by the filter. This leads to the conclusion that, for a given filter at a constant flow rate, the time to reach a fixed headloss depends only on the volume of floc formed from the suspended solids in the raw water and the added chemicals. It should be noted that Eq. 28 implies that the distribution of flocs along the filter depth has no effect on the total headloss. This is in contradiction to many experimental observations showing that deeper penetration of floc results in less overall headloss.

In fact, the determination of the volume of floc removed in the filtration of secondary final effluents would be of little interest and difficult to reproduce. The conventional means of measuring effluent quality is by suspended solids on a mg per ℓ or weight basis.

Tchobanoglous and Eliassen (85) proposed an empirical formula, which fitted their data for the filtration of treated sewage, in the form:

$$H - H_o = k_1(q)^{k_2} \qquad (29)$$

in which q is the amount of deposits, and k_1 and k_2 are constants determined from a log-log graph.

Both works of Mintz (59) and Ives (35) present lists of theories relating to headloss. Not all investigators agree on the exact basis of theory. However, at this stage, the assumption that the increase of headloss is due to the deposits in the filter pores is almost axiomatic.

C. Granular Filters for Wastewater Treatment

Mechanical filtration of primary settled sewage dates back to the year 1883, but passed from favor with the advent of biological methods of treatment. However it regained its popularity in the 1930s due to the development of chemical sewage treatment.

Zack (97) has reported the work done at Wuppertal, Germany; Atlanta, Georgia; York, Nebraska; and San Diego, California, concerned with rapid sand filtration of primary settled effluents. At the Wuppertal plant, rather complete removal of settleable solids at a flow rate of 0.82 gpm/sq. ft. was obtained on a 1 to 2 mm sand filter. Both sand and crushed coal were tried at Atlanta. Using anthracite with an effective size of 0.45 mm, tests indicated filter runs of 12 hr. at rates of 2.0 gpm/sq. ft. and runs of 3 hr. at rates of 3.5 gpm/sq. ft. with less than 5 ft. loss of head. The range of suspended solids delivered to the filter was from 9 to 75 ppm. Comparative data also indicated that twice as many backwashings were required for influents of 57 ppm as for 18 ppm suspended solids.

Streander (83) reported on several installations of deep bed sand filters, among which are Bellefont, Pennsylvania; Grand Canyon, Colorado; and Barrington, New Jersey.

These filters followed trickling filter and activated sludge units. No data were given concerning the operating results of the filters mentioned above. However, he did report, in detail, the design and operating conditions of an experimental pilot filter at Wuppertal, Germany.

Based on the results of the experimental pilot plant, a large-scale installation of sewage filters for treatment of primary settled effluents was installed at Wuppertal in 1939. Ten units each 26.25 ft. wide by 123 ft. long were constructed, giving a treatment capacity of 24 MGD. The filter beds consisted of 28 in. of sand, 1 mm to 2 mm in size, laid over 4 in. of coarse gravel. The probable maximum flow rate was 0.83 gpm/sq. ft., which was based on the pilot plant study. The maximum allowable headloss through the filter was 4.25 ft. The filters removed only about 40 per cent of the SS remaining in the settled sewage. Cleaning of the filters was accomplished with air agitation followed by backwashing with settled sewage. At the Wuppertal plant, the backwash rate was only about 3.7 to 5 gpm/sq. ft. Backwash rates as high as 19 to 25 gpm/sq. ft. expand 1.0 and 2.0 mm sand only 10 per cent. Undoubtedly, an insufficient backwash rate accounted for the low filter efficiency at the Wuppertal plant.

Vosloo (93) made some interesting investigations of rapid sand filtration of final settled effluent without coagulants at the Ancor Sewage Works, Ancor, South Africa. The pressure filter, which was 2 ft. in diameter and 4.5 ft. high, contained 2 ft. of 0.84 to 0.65 mm sand. Several points raised by Vosloo's experiment should be noted: 1) rapid filtration can remove all, or practically all, the suspended solids, at least up to flow rates of 3.33 gpm/sq. ft.; 2) the head required to force the water through the filter at a given rate depended only on the amount of deposits, and not on the flow rate; 3) the filtrability of the wastewater probably depends as much on the nature of the suspended solids as on their quantity; 4) dirty backwash water should be pumped back to the final clarifier, as it has been found that sludge deposited on the filter settle very readily; and 5) with a head of 10 ft. and an average flow rate of 2 gpm/sq. ft., filter run length will range from 6 to 18 hr. if influent solids are around 25 ppm.

Around the 1950s, a considerable amount of work at Luton, England was reported by Pettet, et al. (70-71). Two gravity pilot filters were operated at the Luton plant in parallel: one filter contained 24 in. of sand having a size of 0.85 to 1.7 mm, the other contained 24 in. of anthracite having a size of 1 to 2 mm. A flow rate of 2 gpm/sq. ft. was used and the allowable headloss build-up was 7 ft. The filters were backwashed once every 24 hr. using 2 to 3 percent of the filtrate produced during the previous run. Backwashing was preceded by air agitation of the media, and backwash rates were 13.3 and 10.0 gpm/sq. ft. for the sand and anthracite respectively. The results showed almost identical SS (86 per cent) and BOD (63 per cent) removal efficiencies for sand and anthracite filters. A three month test period on two filters using the same sand as before with depths of 2 and 3.5 ft. showed that there was no significant difference in effluent quality until the flow rate exceeded about 4.5 gpm/sq. ft., at which point the deeper bed gave slightly better results. Pettet, *et al.* (71) argued that it was doubtful that this slight improvement at high rates of filtration would justify using the extra depth of medium on a full-scale plant, especially since the power required for pumping would also be increased. It was thought that the greater depth might prevent a "break-through" of solids, but during experiments with a medium 2 ft. deep, the maximum loss of head was reached almost invariably before a break-through occurred. On the basis of preliminary and pilot plant tests, a full-scale plant with 6 rapid sand filters capable of treating up to 7.2 MGD was put into operation at Luton in 1951 (17-19).

Nicolle (64-65) reported the work at the Pretoria sewage-treatment works, Johannesburg, South Africa, in which five rapid sand filters were designed to treat 3.0 MGD of secondary settled effluent at a flow rate of 3 gpm/sq. ft.

under a maximum head of 9.0 ft. He reported that filters with a sand size of 0.84 to 0.59 mm gave an effluent with an average suspended solids of 5 mg/ℓ. With sand of 1.4 to 0.65 mm, the suspended solids in the effluent averaged 8 mg/ℓ. With a headloss termination point of 6.5 ft., run length varied from 8 to 10 hr. Backwashing was preceded by air wash. At a backwashing rate of 25.7 gpm/sq. ft., wash-water used was about 9.5 per cent of the filtrate of the previous run.

Work concerning tertiary treatment of secondary effluents to obtain a water for injection into coastal sea water intrusion barriers has been done at Los Angeles. Laverty, et al. (45) reported on six months of preliminary tests which were made with a rapid sand filter at the Los Angeles Hyperion treatment plant. Standard-rate activated sludge plant effluent was applied at a rate of 2.0 gpm/sq. ft. to the filter which was an 11 in. bed of sand having an effective size of 0.95 mm and a uniformity coefficient of 1.6. Removal efficiency for suspended solids was 46 per cent, for BOD was 51 per cent.

Fall and Kraus (21) reported on a laboratory study of tertiary treatment for high-rate, activated sludge effluent at the Peoria Sanitary District Treatment Works. Their results showed a 50 per cent reduction in suspended solids and a 40 per cent reduction in BOD at a flow rate of

1.12 gpm/sq. ft. The average filter influent suspended solids and BOD were 16.0 and 28.3 mg/ℓ respectively. The filter clogged readily and run lengths were only a few hours long. The characteristics of the filter media were not mentioned.

Merry (57) conducted a study at the Pollution Control Plant, Ames, Iowa, to investigate the application of rapid sand filtration as a means of tertiary treatment of final settled effluent from a standard rate trickling filter plant. The pressure filters had a 24 in. depth of sand with an effective size of 0.55 mm and a uniformity coefficient of 2.36. Runs were terminated when the head reached the allowable 6.25 ft. Average suspended solids removal efficiencies were 71.2, 67.6 and 64.4 per cent when flow rates were 2, 4 and 6 gpm/sq. ft., respectively. Average BOD removal efficiencies were 56.8, 55.5 and 51.6 per cent, when flow rates were 2, 4 and 6 gpm/sq. ft., respectively. Run lengths were 10, 6.25 and 4.67 hr. for flow rates of 2, 4 and 6 gpm/sq. ft., respectively. No air wash was used during backwashing. A backwash rate of 20 to 25 gpm/sq. ft. was used for a period of 10 minutes.

The first mixed-media filtration for polishing the extended aeration effluent at a municipal sewage treatment plant, Philomath, Oregon, was reported by Culp and Hansen (12). The filtration rate was 5 gpm/sq. ft. Rather high removal efficiencies both in suspended solids and BOD (85-99 per cent) were reported. The turbidity of the filtrate was consistently less than the USPHS drinking water standard of 5 JTU and was as low as 0.3 JTU. No information concerning the media size and grade, terminal headloss and run length was reported.

Recent developments on the rapid sand filters at Luton, England were reported by Naylor, et al. (63). A comparison between the gravity filter and up-flow filter was made. Filter media used for both downward-flow and upward-flow filters was a single grade of sand, 0.85 to 1.7 mm. The downward-flow filter with media depth of 36 in. operated at a flow rate of 4 gpm/sq. ft. resulted in a suspended solids removal efficiency of 41 per cent and BOD removal efficiency of 42 per cent. An upward-flow filter with media depth of 60 in. operated at varying rates with an average of 3.92 gpm/ sq. ft. resulted in a suspended solids removal efficiency of 56 per cent and a BOD removal of 51 per cent. They concluded that the upward-flow filter consistently produced a better quality effluent even at dosing rates approximately 50 per cent higher.

In the discussion of the paper of Naylor, et al., Holden (24) questioned whether the increase in the depth of sand in the downward-flow filter from 24 in. to 36 in. had resulted in improved performance. Holden's experience at

Cambridge, England, with a pilot-scale filter had shown that nearly all of the solids were removed in the top 12 in. of the bed and that after a normal run the sand was fairly clean 12 in. below the surface. For the filter plant at Cambridge, which had a media size similar to that of Luton, the optimum flow rate for both filter types was 2.5 gpm/sq. ft. with a feed containing 20 to 40 mg/ℓ SS. At this rate, the run length for the downward-flow filter was 6 hr. and for the upward-flow filter was 24 hr.

American engineers have also been active in designing and using granular filters as a means of tertiary wastewater treatment, particularly during the last five years. Berg and Brunner (2) reported on the tertiary filter plant at Lebanon, Ohio, in which secondary effluent from a conventional activated sludge plant received further treatment by pressure filters. Two different filter beds were evaluated: a single medium filter consisting of 20 in. anthracite coal having an E.S. of 0.75 mm and U.C. of 1.5 and a dual-media filter consisting of 6 in. of sand and 14 in. of coal. The E.S. of the sand was 0.45 mm and the U.C. was 1.6. Filter runs with both single and dual media were made at rates of 5 and 10 gpm/sq. ft. The effect of polyelectrolyte on filter performance was evaluated by using several polyelectrolytes (two cationic, one anionic, and one nonionic) with a single-medium filter and one cationic polyelectrolyte and alum with a dual media filter.

The nonionic and anionic polyelectrolytes were less effective in preventing floc breakthrough than the cationic. The optimum chemical dosages were 12.5 mg/ℓ of alum and 2.5 mg/ℓ of cationic polyelectrolyte. Under this condition filters produced a good filtrate with a reasonable rate of headloss increase. The presence in the suspended solids of certain organisms or soluble materials as well as particle size and Zeta potential are all factors that can alter floc penetration and subsequently filter run length. The dual media and single medium run lengths compared closely with comparative influent suspended solids, which suggested filtration takes place in the upper layers of the media as opposed to penetration through the bed.

However, this writer disagrees with Berg and Brunner's conclusion. The size of coal (E.S. 0.75 mm, U.C. 1.5) used in their dual media filter was not a proper one, since there was, in fact, a size of less than 0.7 mm coal in the top few inches of the filter bed, a size too fine to provide the penetration of solids to a significant depth. Most of the bed, except the top few inches, was not utilized optimally. A proper selection of coal size and its gradation, such as using a coarser anthracite size with a smaller size range, enhances the appeal of dual media filtration.

Rapid sand filtration for tertiary wastewater treatment at the Metropolitan Sanitary District of Greater Chicago (54-55, 78) is worth noting. The Hardinge sand filter utilizes a silica sand with an effective size of 0.51 mm and a uniformity coefficient of 1.62. The Hardinge type filter utilizes a traveling backwash carriage, carrying over each cell module a hooded assembly. This traveling arrangement allows the filter to remain in continuous service while segmentally washing a single cell.

Poor correlation was obtained between effluent quality and flow rate, effluent quality and solids loading, and solids removal and flow rate. However, Lynam, et al. (55) found that there was a degree of correlation between the suspended solids removal and suspended solids applied to the filter. They also found that the filter effluent quality depended on the secondary effluent quality in combination with flow rate.

A pilot plant investigation of sand filters for tertiary wastewater treatment at Derby, England, has been reported by Oakley and Cripps (66) and Joslin and Greene (39). Information concerning detailed descriptions of the physical characteristics of these filters and their results and operational data will be directed to the aforementioned papers. Average suspended solids removal was 47 to 76 per cent with no apparent correlation between removal and filtration rate. BOD removals are reported as 21 to 69 per cent. It was concluded that a graded 1.0 to 2.0 mm sand was most satisfactory at Derby and that there was little difference in the performance of downward-flow and upward-flow units.

The formation of "mud balls" and a coating of solids on the filter walls were observed. This might be due to insufficient backwashing, since these filters were backwashed at a rate just sufficient to fluidize all the filter medium, and air scour was not used.

High-rate granular media filtration also finds its application in the area of industrial wastes. Among the largest industrial water users are the steel manufacturers. Donovan (13) presents a detailed study of filtration of steel mill wastes in which the primary pollution is suspended solids. A pilot plant study includes filter runs with single anthracite medium, single sand medium, and then the optimum combination of these two media into anthracite-sand dual media filters. A full-scale plant was designed based on the results of pilot-scale tests in which the filters consisted of 24 inches of 2 to 3 mm sand under 60 inches of 5.1 mm anthracite. These filters were designed to be operated at a rate of 16 to 18 gpm/sq. ft. to produce an effluent of less than 15 mg/ℓ suspended solids. Cationic polyelectrolyte was applied as a chemical coagulant. During backwashing, the filter was air-scoured for 5 minutes at

8 cfm/sq. ft. then flushed with water at 30 gpm/sq. ft. for 10 minutes. This procedure proved satisfactory and no agglomerates were found in the full-scale filters after a year's operation. Backwashing the filters at regular 12-hour intervals resulted in a backwash water use of about 2 to 3 per cent of the filtrate from the previous run. The material washed from the filter still remained in fairly agglomerated form, particularly if a polyelectrolyte was used. However, it clarified rapidly by simple sedimentation.

Probably the most extensive study in the area of tertiary wastewater filtration was the recent work of Tchobanoglous (84) and Tchobanoglous and Eliassen (85), both at Stanford University. Their studies attempted to: 1) understand the variables which control the process; 2) determine the pertinent mechanism or mechanisms responsible for the removal of particulate matter from a wastewater; and 3) develop equations which can be used to describe the time-space variation and build-up of material within the filter.

Settled sewage effluent used in their study was obtained from a pre-assembled "Rapid Bloc" type of pilot-scale activated sludge treatment unit. Influent characteristics which were evaluated included the relationship between turbidity and suspended solids, particle size and distribution, and particle charge and distribution.

As for the investigation of filtration process variables, they investigated the effect of filtration rate, sand size and the development of headloss with time. They found that in the absence of chemical coagulants removal of suspended solids is primarily a function of the filter medium grain size. Thus, at a flow rate of 2 gpm/sq. ft. the SS removal efficiency varied from 40 per cent for 0.5 mm effective size sand to 15 per cent for 1.0 mm effective size sand. These low removal rates are presumably due to the nature of the activated-sludge solids and to the relatively low SS concentrations of 10 to 20 mg/ℓ in the final settled effluent.

Another study of tertiary wastewater filtration has been carried out by Tebbutt (86) at the University of Birmingham. Effluent from the Minworth works of the Upper Tame Main Drainage Authority was applied to both laboratory-scale filters and in-plant filters. His investigations were focused primarily on the media size and flow rate effects on the SS, BOD, and COD removal efficiencies.

Removal of suspended solids from the Minworth effluent ranged from 38 to 70 per cent during the tests and it would appear that the largest sand used (2.4 to 4.7 mm) was significantly less effective than the medium sized sand (1.2 to 2.4 mm). Fine sand (0.5 to 1.0 mm) and a dual media bed offered no improvement in SS removal over a 1.0 to 2.5 mm anthracite bed. For all the media examined at a fixed depth of 24 in., the flow rate did not affect the removal of suspended solids over the range of 1 to 5 gpm/sq. ft.

In series tests, there was no significant difference between the removal efficiency for BOD (66 to 71 per cent) achieved by the different filters although the fine sand filter might possibly be slightly more effective in removing COD (36 to 48 per cent), perhaps because of the greater adsorptive capacity of the larger surface area in the bed.

Tebbutt recommends that laboratory or pilot scale studies be carried out over a period of at least 12 months on any proposed tertiary wastewater filter installation due to the random nature of suspended solids in the effluent from final clarifiers. Appendix A is a summary of published data concerning granular filters for wastewater filtration.

D. A Summary

The essential conclusions of this literature review can be summarized as follows:

(1) Considerable work has been done which has led to a better understanding of the granular filtration process and produced results useful for practical application. However, the basic removal mechanisms have not been supported unequivocally by experimental or theoretical work.

(2) The mathematical models are still restricted by idealized hypothetical assumptions. Therefore,

their practical applications are limited.

(3) Most designs of wastewater filtration systems are based on the past experience of water filtration. The significant difference in influent characteristics between water and wastewater has not been fully recognized and understood. Thus, this basic difference has not been incorporated into the design of wastewater filters.

(4) Dual-media or multi-media filters are essential in order that depth removal occur in wastewater filtration due to easy clogging in the top layers. A rather narrowed range of coal size is recommended.

(5) Air agitation or a mechanical device to break up mud balls prior to water backwashing is required to guarantee a sufficient backwash. Insufficient backwash results in growths inside the filter and poor effluent quality.

(6) No optimization concept has yet been applied to wastewater filtration.

The use of granular filters in tertiary wastewater treatment is advocated even though much basic knowledge about the design and operation of such filters in wastewater treatment is unknown.

IV. EXPERIMENTAL INVESTIGATIONS

A. Basis for Experimental Pilot Plant Design

Two obvious facts concerning wastewater filtration which were revealed in the literature review in the previous chapter (Chapter III) are:

(1) surface clogging resulting in a high rate of headloss build-up and short filter runs predominates in wastewater filtration, and

(2) no mathematical model, which takes into consideration surface removal mechanisms is available to predict filter performance.

Therefore, a modified design of filter media to reduce or eliminate surface clogging and a pilot plant design which is able to investigate filter behavior in each subsequent layer of the filter bed are two important subjects which deserve considerable attention.

1. Type and size gradation of filter media

Surface clogging during wastewater filtration can be reduced by: 1) using a narrowed range of medium size, preferably uni-sized media; 2) using dual or multimedia filters, 3) using a coarse to fine flow direction (upflow) in a graded media filter.

The range in media size is quite important; a filter with a small range of sand size (low U.C.) provides longer run lengths (83). How narrow this sand size range should be has not yet been definitely determined. The best size range is influenced, to some extent, by the pretreatment given the final effluent (chemical precipitation) and the amount of solids to be removed.

Uni-sized media filters have been operated and investigated primarily on a research basis (25, 26, 30-34, 47, 84-85). The reason for using a uniform grain size in the research was the simplicity of predicting the filter performance. A systematic study of iron removal comparing the filter performance of graded media and uni-sized media was conducted by Kim in Alaska (40). He claimed that a properly designed uni-sized media could increase the run length more than ten times over that of graded sand filters of the same effective size. Apparently, uni-sized sand reduced considerably or eliminated the surface clogging problem.

By using uni-sized media, the cost of media is expected to increase. However, the portion of filtration cost contributed by the filter media is only a small portion of the total cost of a filter plant. Therefore, it is the writer's opinion that the added cost of a uni-sized media should not be used as an argument against its use.

A second approach to reduce surface clogging is through the use of dual or multi-media filters. Conley (10) advocated the use of a dual media filter composed of coarse anthracite on top of finer sand to attain filtration in the direction of diminishing grain size. Ives (31) reported some results of triple-media filtration, in which a third material, garnet sand, was installed beneath the silica sand. Mohanka (62) even tested a five-layer filter using polystyrene, anthracite, sand, garnet, and magnetite located from top to bottom in the order listed. Triple- and five-media filters showed, respectively, 33-67 per cent and 37-60 per cent of the headloss of a sand bed of equal depth. In all these modifications, the fine grain at the top of each medium cannot be avoided if graded materials are used. Also, the use of a very heavy, small grain at the bottom requires extreme care in the composition of the supporting gravel, lest underdrain clogging occur. Thus, a further modification to dual media filters should be made by using as close to uni-sized material as possible in each layer.

It should be noted that a uniform size of medium is impossible to achieve in practice. However, a medium graded into the narrowest size range of available sieves should result in a nearly uniform sized medium.

Therefore, both a uni-sized sand filter and uni-sized anthracite and sand dual media filters were adopted in this study.

2. Pilot plant design

The conventional way to study filter removal behavior has been to use a series of sampling points within the depth of a single filter. This technique is subject to certain serious objections. First, unless the sampling points are well designed they may interfere with the structure of the bed and, thus, the flow pattern in the filter may be altered and give anomalous results. Second, the withdrawal of the sample disturbs the flow and may disturb the deposited matter, which results in a nonrepresentative sample being collected. Third, it is also apparent that the flow through the filter is reduced at every sampling point by the sample drawn off, and corrections have to be made for this.

In order to investigate filter behavior in each subsequent layer of a filter bed, three filter sets each containing four filter cells (total of 12 filters) were constructed. Each filter set consisted of four filters containing media of different depths and each filter was equipped with its own pressure regulator and effluent rate control. This arrangement was similar to the apparatus used by Ison and Ives (30) and Hsiung and Cleasby (26) and was designed to overcome objections to the methods of sampling previously employed in most filtration studies.

The effluent of each filter in the 4-depth filter set indicates the water quality which can be expected after filtration through the corresponding depth of filter media. This information is assumed to be equivalent to information obtained from a single deep filter with multiple sampling outlets at the depths corresponding to those used in the 4-depth filter set. This is true as long as the media in each filter is in the same condition and subjected to the same influent suspended solids and flow rate. The advantages and disadvantages of the 4-depth filter set have been listed by Hsiung (25). It should be noted that for filter beds of thin depth the method of media placing and the presence of the retaining stainless steel bottom and walls result in variations in bed structure both transversely and in depth. Since the rate of change of the suspended solids concentration is maximum at the filter bed surface, any such variations in a thin bed will result in anomalous values of the effluent quality.

B. Pilot Plant Description

In order to study filtration of wastewater for developing optimization procedures, a pilot plant was constructed and operated at the Ames Pollution Control Plant, a standard-rate trickling filter plant with an average flow of 6.0 MGD. The wastewater entering the plant consists primarily of domestic wastes from the City of Ames and Iowa State University. A number of small industries, including the

National Animal Disease Laboratory and Hach Chemical Co., also contribute to the waste load. Overall plant BOD removal efficiencies range from 75-80 per cent in winter to about 90 per cent in summer. The suspended solids concentration in the final settled effluent, which was used as influent to the pilot plant filters, ranges from 20 to 40 mg/ℓ in winter and about 10 to 20 mg/ℓ in summer. The BOD in the final settled effluent ranges from 30 to 60 mg/ℓ in winter and from 15 to 25 mg/ℓ in summer.

In Ames, there is a fairly typical variation in the characteristics of final settled effluent during a day, the highest strength of final settled effluent occurring at the noon hour, and the lowest strength occurring about 8 a.m. Fig. 1 shows the typical variation in turbidity, SS and BOD.

A flow diagram of the pilot plant is shown in Fig. 2. The pilot plant was constructed inside a building which was originally built to house the chlorinators near the manhole at which the effluent from the three final clarifiers are joined together. Because of the constant mixing and churning action of the effluent in this manhole, wastewater withdrawn from here to supply the pilot plant filter is representative of the plant effluent.

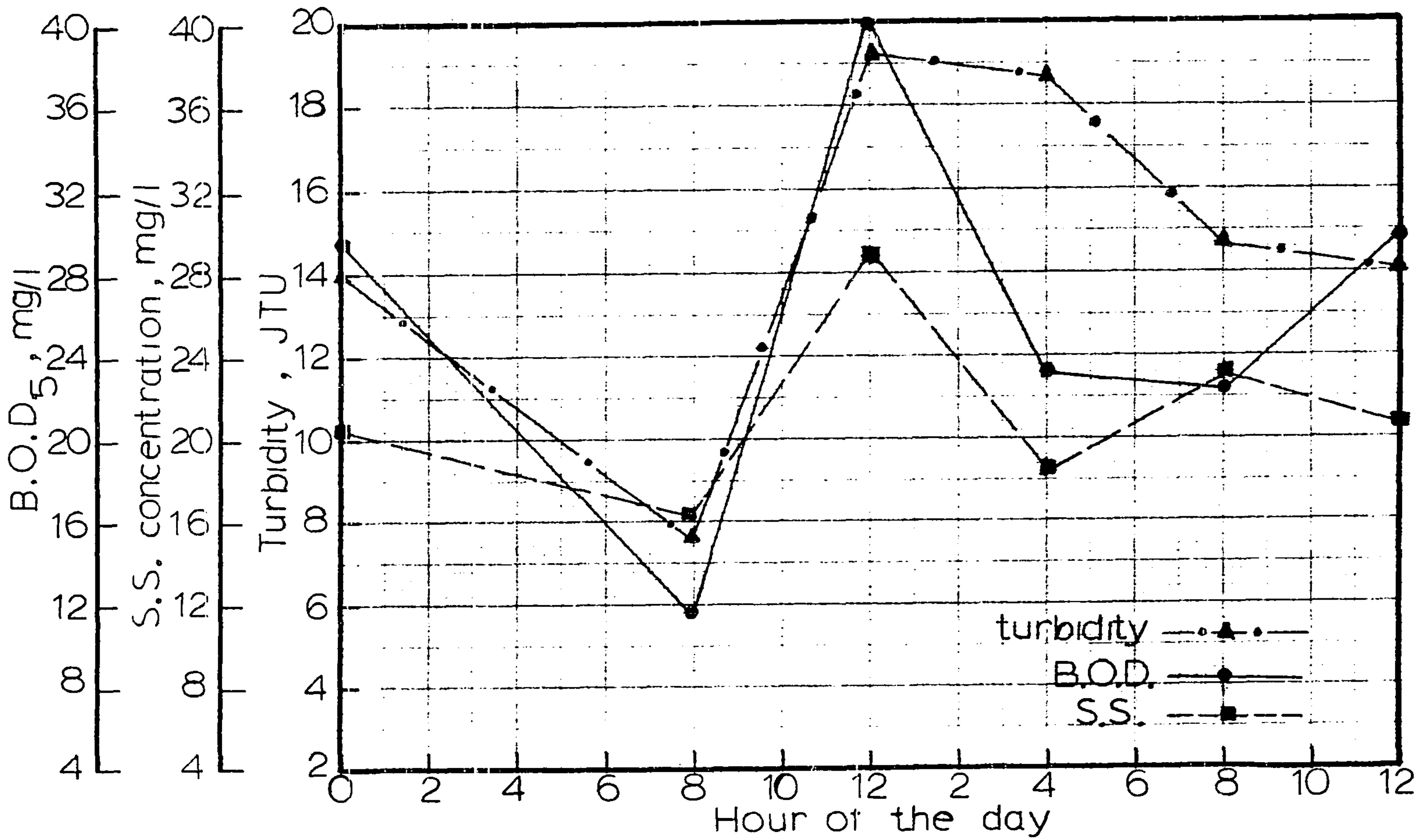

Fig. 1. Typical final effluent characteristics at Ames Pollution Control Plant. May 3, 1971

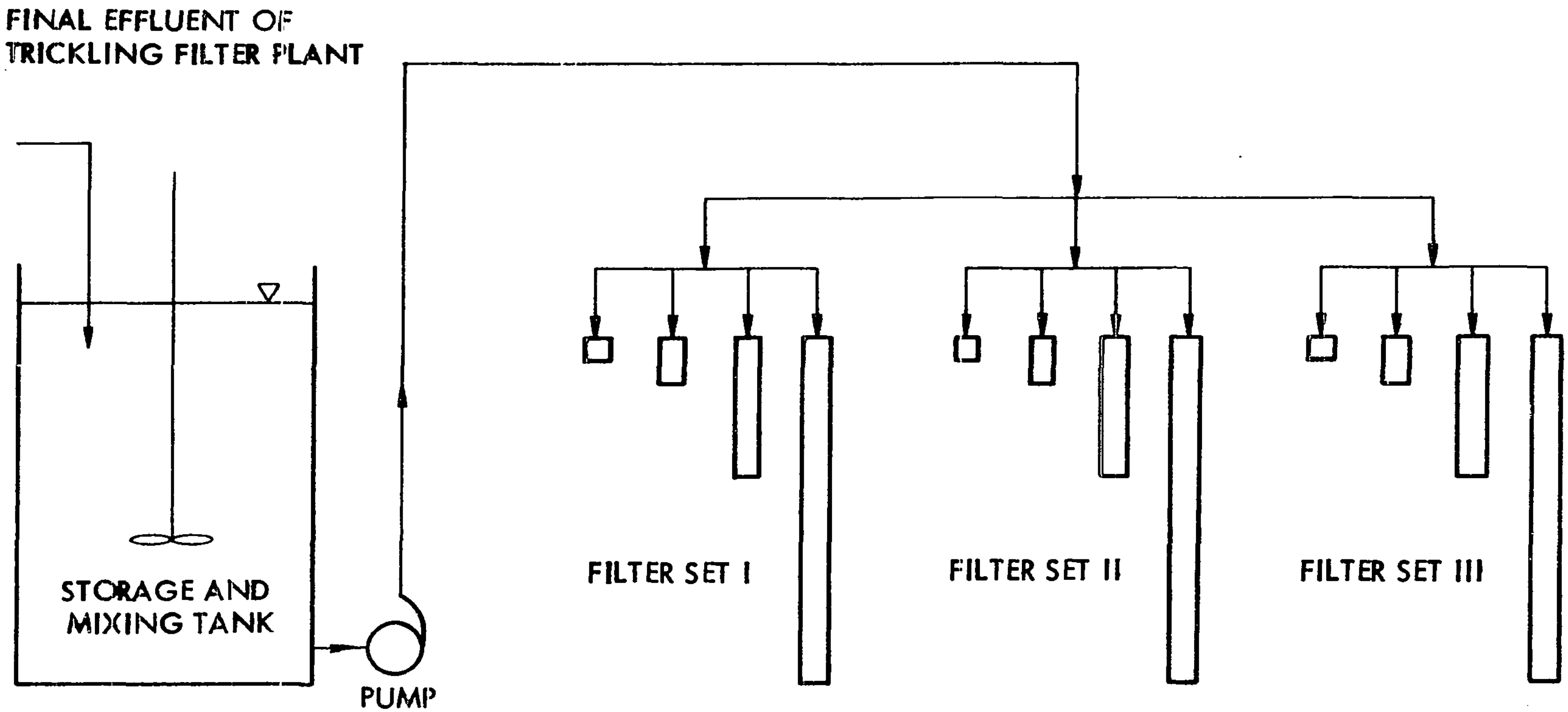

Fig. 2. Schematic arrangement of pilot tertiary wastewater filters

C. Experimental Apparatus

1. Filter apparatus

The filter apparatus consisted of three filter sets each containing four filter cells (total of 12 filter cells). Each filter set consisted of four filters of different depths and each filter was equipped with its own pressure regulator and effluent rate control, as shown in Figs. 3a, 3b and 3c.

The various components of the filter apparatus are described in the following paragraphs.

a. The filter housings   The filter housings were constructed of 4 in. inside diameter plexiglass. This provided a filter area of 0.0873 sq ft. The four filter cells can provide media depths of 1, 5, 14, and 24 in., respectively. A splash plate was located at the top of each of the filter housings to prevent the incoming flow from disturbing the media surface. The underdrain system of each housing consisted of a layer of U.S. Standard #50 stainless steel mesh, strengthened underneath with a U.S. Standard #10 mesh layer so that a flat bottom was obtained.

b. Filter media   Two kinds of filter media were used in this study: anthracite coal and sand. The anthracite coal used in this research was obtained from the Reading Anthracite Coal Company, Pottsville, Pennsylvania. It was designated as "Philterkol". The type of coal used is typical of the anthracite coal used frequently for filter media in

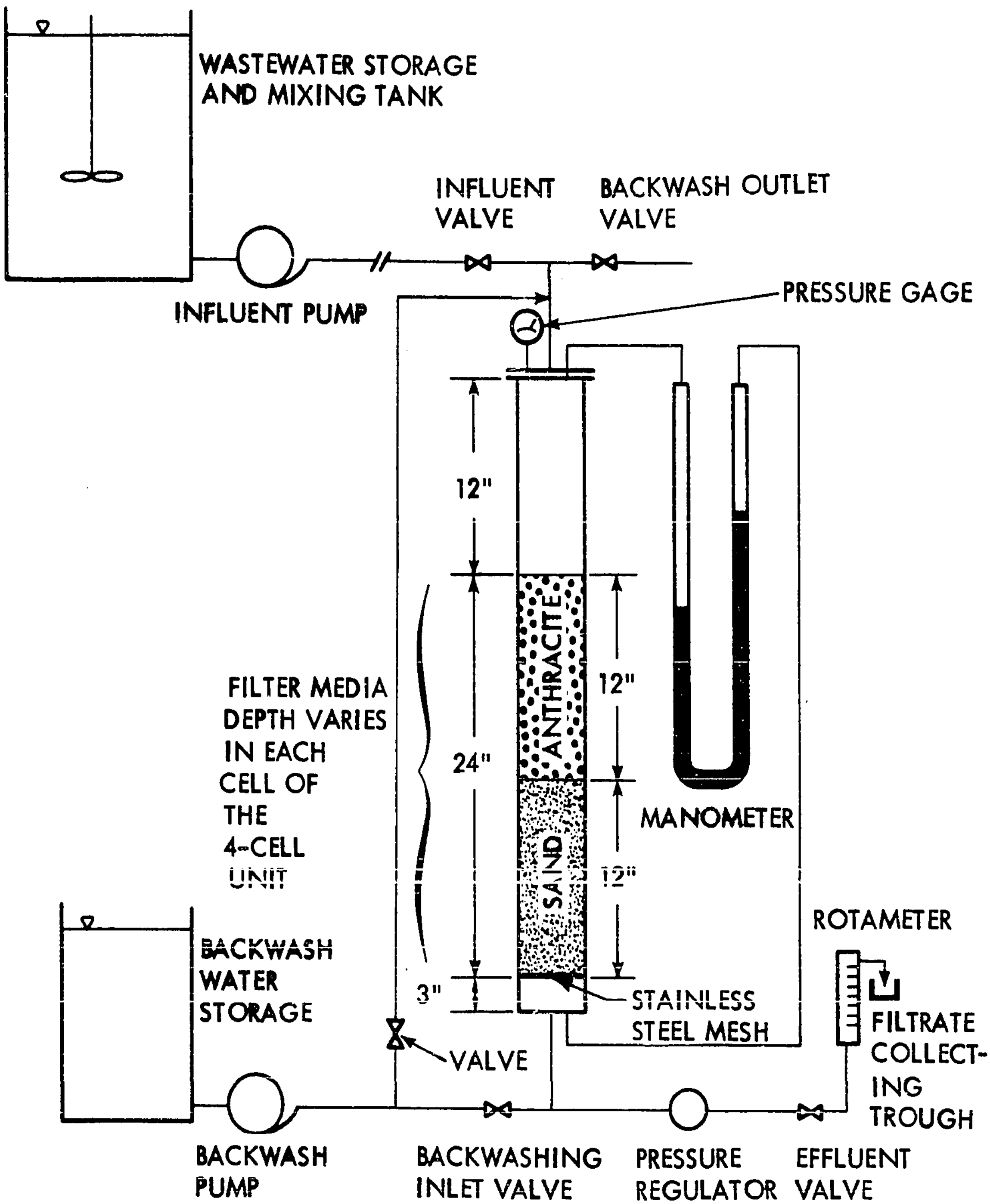

Fig. 3a. Schematic diagram of individual filter cell (four cells in each filter set)

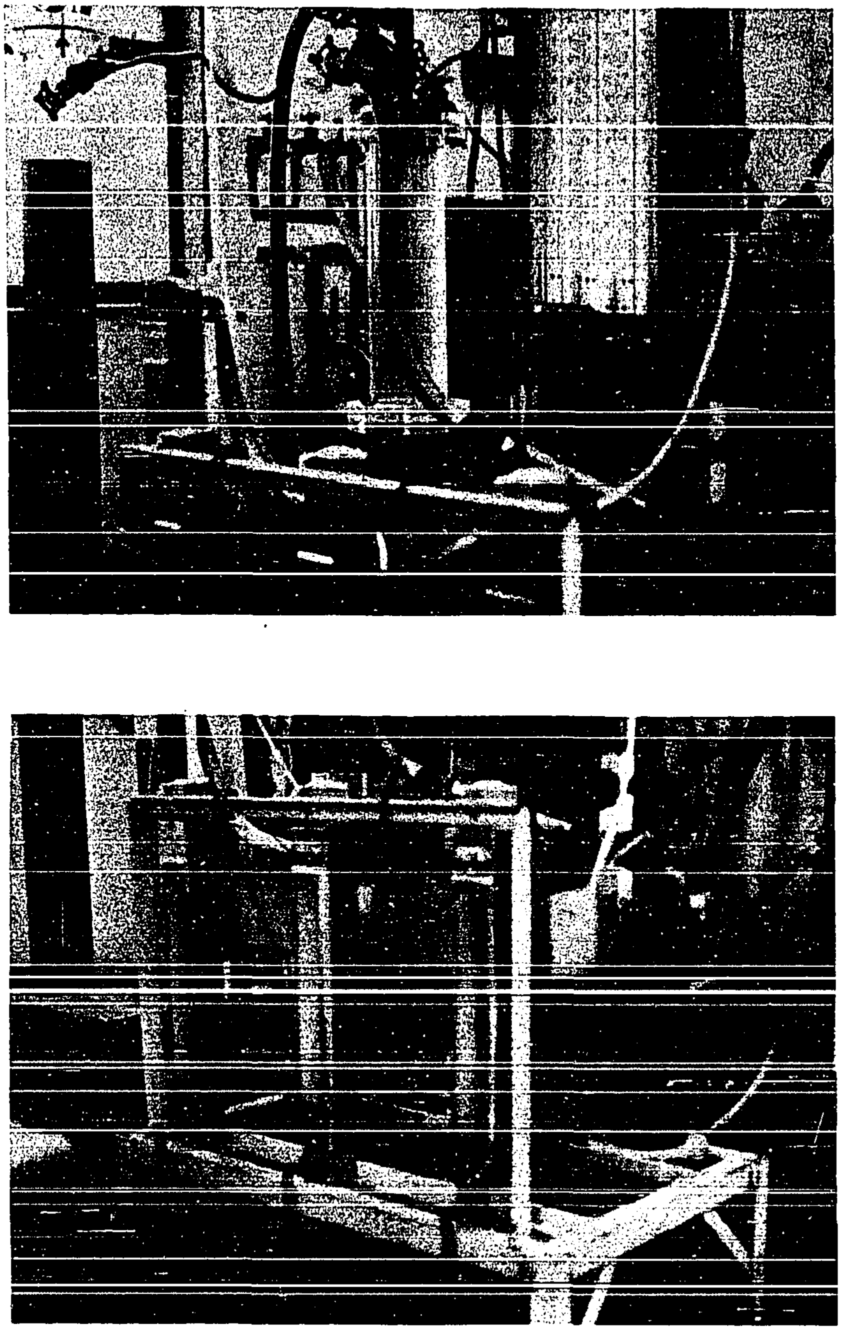

Fig. 3b. Pilot filter apparatus.

Fig. 3c. Pilot filter apparatus.

single, dual and multi-media filters. Two types of coal with different grading were purchased: one with an effective size of 1.00 mm and a uniformity coefficient of 1.8, the other with an effective size of 1.10 mm and a uniformity coefficient of 1.65. The uni-sized anthracite used throughout this project was prepared by sieving the "Philterkol" anthracite in a Gilman set of U.S. Standard sieves on a mechanical shaker for 5 minutes.

The sand used was a granular filter sand obtained from the Northern Gravel Co., Muscatine, Iowa. The uni-sized sand used in this study was prepared by sieving it in a Gilman set of U.S. Standard sieves on a mechanical shaker for 10 minutes.

The uni-sized sand or anthracite used was that media 100 per cent of which passed one sieve number and was retained on the next sieve number. For example, a 10/12 anthracite was one in which 100 per cent of the material passed a U.S. Sieve No. 10 and 100 per cent was retained on a U.S. Sieve No. 12. A screen analysis of commercially graded anthracite,[1] from which a uni-sized 1.84 mm anthracite (passing U.S. Sieve No. 10 and retained on U.S. Sieve No. 12) was obtained, is shown in Table 1a.

Table 1b shows the size range of uni-sized sand and anthracite used for this study.

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[1]Philterkol Special No. 1, Reading Anthracite Coal Company, Pottsville, Pennsylvania.

Table 1a. Screen analysis of Philterkol Special No. 1*

Openings Inches	Milli-meters	Tyler Mesh	U.S. No.	Per Cent Weight Retained	Per Cent Cumulative Weights
.132	3.36	6	6		
.0937	2.33	8	8	3.0	3.0
.0787	2.00	9	10	19.4	22.4
.0661	1.68	10	12	22.9	45.3
.0555	1.41	12	14	22.7	68.0
.0469	1.19	14	16	14.0	82.0
.0394	1.00	16	18	9.8	91.8
.0331	0.841	20	20	5.2	97.0
.0278	0.707	24	25	2.2	99.2
.0234	0.595	28	30	0.2	99.4
.0165	0.420	35	40	0.2	99.6
.0117	0.297	48	50	0.2	99.8
		-48	-50	0.2	100.0

* E.S. 1.00 to 1.10 mm, U.C. max 1.80.

Table 1b. Size range of uni-sized sand or anthracite

Geometric mean size mm	Sieve opening		U.S. sieve No.	
	Passing mm	Retained mm	Passing	Retained
0.46	0.50	0.42	35	40
0.55	0.60	0.50	30	35
0.65	0.71	0.60	25	30
0.77	0.84	0.71	20	25
0.92	1.00	0.84	18	20
1.09	1.19	1.00	16	18
1.30	1.41	1.19	14	16
1.54	1.68	1.41	12	14
1.84	2.00	1.68	10	12

c. Manometer In each filter set, a multiple tube manometer[1] was provided for measuring the headloss build-up in each filter cell. A 1/4 in. copper pipe was connected from the top of the filter housing to the bottom end of the manometer, while another 1/4 in. copper pipe was connected from the bottom of the filter housing to the top end of the manometer. The manometer is based on the principle of a U-tube, which measures the pressure differences. Two kinds of indicating fluids were used: mercury (specific gravity 13.6) and red Merian fluid (specific gravity 2.95). Mercury was used for the third and fourth filter cells of each filter set, in which media depths were 14 and 24 in. respectively. One inch of mercury shown in the manometer indicated that the leadloss across that filter media was 1.05 ft. of a water column. Red merian fluid was used for the first and second filter cells, in which media depth were about 1 and 5 in. respectively. One inch of red merian fluid head in the manometer indicates that the headloss across that filter media was 0.163 ft. of water column. By using red merian fluid small increments of head loss in shallow filter cells could be detected.

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[1]Model 33KB35, Multiple Tube Manometer, Meriam Instrument Division, The Scott & Fetzer Company, Cleveland, Ohio.

d. Rotameter The effluent from each of the filters passed through a variable area, float type, flow meter.[1]

The flow meters were capable of measuring flow from 0.1 to 0.8 gpm. The flow meters were calibrated initially by collecting a timed sample in a two-liter volumetric flask. During operation of the apparatus, a small deposit would accumulate on the inside of the glass tube and on the float of the flow meters. These deposits were removed after each run by flushing the flow meters with clean water. Occasionally, it was required to take the rotameters apart and clean them manually with soap and water.

e. Pressure regulator A Type 95 Pressure Regulator[2] which reduces a pressure range of 15 - 300 psi to 5 psi was installed in front of the rotameter of each filter cell. The pressure regulator maintained a constant pressure on the rotameter regardless of the pressure change occurring in the filter housing due to clogging inside the filter pores. Thus, a constant flow rate was maintained throughout a filter run.

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[1] Model 1112A, Full-View (O-Ring Seal) Rotameter, Brooks Instrument Division, Hatfield, Pennsylvania.

[2] Fisher Governor Company, Marshalltown, Iowa.

2. Turbidimeter and Millipore filter

A Hach turbidimeter[1] was used for monitoring both influent and effluent wastewater turbidity. A true nephelometer, the turbidimeter operates on the principle of measuring scattered light. A sensitive photomultiplier tube gives instant response by converting the reflected light to an electrical signal which is measured on the panel meter. Turbidity ranges are selected electrically. Turbidity is read directly in five different ranges on a single scale card (0 - 0.1, 0 - 1.0, 0 - 10.0, 0 - 100 and 0 - 1000 scales) inserted in the meter face.

Suspended solids in the influent and effluent water were collected using the Millipore filter.[2] For the influent wastewater, 100 ml samples were filtered through a filter glass pad, and the suspended solids collected on the pad were determined according to Standard Methods (79). A sample size of 200 ml was used for the filtrate, since only a negligible amount of suspended solids existed in the filtered samples. SS concentrations were determined from the sample size used and solids recovered.

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[1]Model 2100, Hach Chemical Company, Ames, Iowa.

[2]Millipore Filter Corporation, Bedford, Massachusetts.

3. <u>Pumps, wastewater storage and mixing tank, and clean water storage tank</u>

A total of three pumps were used in this study. The first pump was used to withdraw wastewater from the plant outfall sewer and deliver it to the wastewater storage and mixing tank which had a capacity of 4,000 gallons. Once the storage tank was filled, its contents were continuously mixed using a 14 in. diameter propeller mixer. A second pump was used to pump the uniformly dispersed wastewater to the filter units. The third pump was used to pump clean water, which was stored in a 100 gallon backwash water storage tank, to the units for backwash purposes. The clean water was obtained by filtering the normal plant water supply through an activated carbon filter.[1] The water source contained a high iron content, about 8-9 mg/ℓ, and was filtered to remove the iron.

D. Operation of Pilot Plant Apparatus During a Filter Run

1. <u>Measurement of influent and effluent wastewater quality</u>

a. <u>Turbidity and suspended solids</u> The influent and effluent wastewater qualities were evaluated on the basis of their suspended solids content. However, due to the diffi-

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[1]Everpore, Inc., Oak Brook, Illinois.

culty in making an SS test for each sample collected during a filter run a correlation was made between sample turbidity and suspended solids content. Then, direct readings of turbidity on each sample were made because of the ease with which tests could be performed in the field.

It should be understood that turbidity is an expression of the optical property of a sample which causes light to be scattered and absorbed rather than transmitted in a straight line through the sample. Special care needs to be exercised if an attempt is to be made to correlate turbidity with suspended solids, as the size, shape, and refractive index of the particular materials are most important optically but bear little direct relationship to the concentration of the suspended solids.

Therefore, equations relating the turbidity and suspended solids were revised periodically. It was found that the same sample with one suspended solids content had a higher turbidity reading on the 0-10 scale and a lower turbidity reading on the 0-100 scale of the Hach turbidimeter, due to the difference in sensitivity of the machine. For example, one sample having an SS of 9.0 mg/ℓ read 10 JTU on the 0-10 scale, but read 8.5 JTU on the 0-100 scale. Thus, two standard equations relating the turbidity and SS content were developed from more than sixty samples each time: one accounts for a sample with low turbidity when reading the turbidity on the 0-10 scale; the other accounts for a sample with high turbidity when reading the turbidity on the 0-100 scale. These equations are summarized in Table 2 and shown graphically in Fig. 4.

As indicated in Table 2, a reasonably good relationship could be established between the field turbidity readings and the sample suspended solids. During each run, 12 samples (3 from the influent and the rest from the effluent of the fourth filter cell of each set, media depth 24 in.) were measured for both turbidity and SS. The correlations from each run were developed and checked against that of the standard correlation equations. It was found that it checked closely with the standard correlation equations.

All readings in turbidity were converted using the proper standard equation into suspended solids for further analysis.

b. Other chemical tests Other chemical tests made to evaluate filter influent and effluent qualities included BOD, COD, TOC, orthophosphate, total phosphate, organic nitrogen, ammonia, nitrate and nitrite. These determinations were made according to Standard Methods (79). However, complete chemical tests of influent and effluent samples were made in only a few typical runs in which composite samples of influent and effluent were collected throughout the filter run.

Table 2. Turbidity-suspended solids relationships

Date	0-10 Scale	Correlation Coeff., %	0-100 Scale	Correlation Coeff., %
Dec. 23, 1970	SS[a] = -0.18 + 0.72 (JTU)[b]	82.95	SS = -0.92 + 1.38 (JTU)	92.65
Apr. 2, 1971	SS = -1.34 + 0.93 (JTU)	87.52	SS = -5.53 + 2.32 (JTU)	98.93
June 14, 1971	SS = -1.48 + 0.65 (JTU)	76.24	SS = -0.62 + 1.60 (JTU)	89.63

[a]SS = Suspended solids concentration in mg/ℓ.

[b]JTU = Jackson Turbidity Unit.

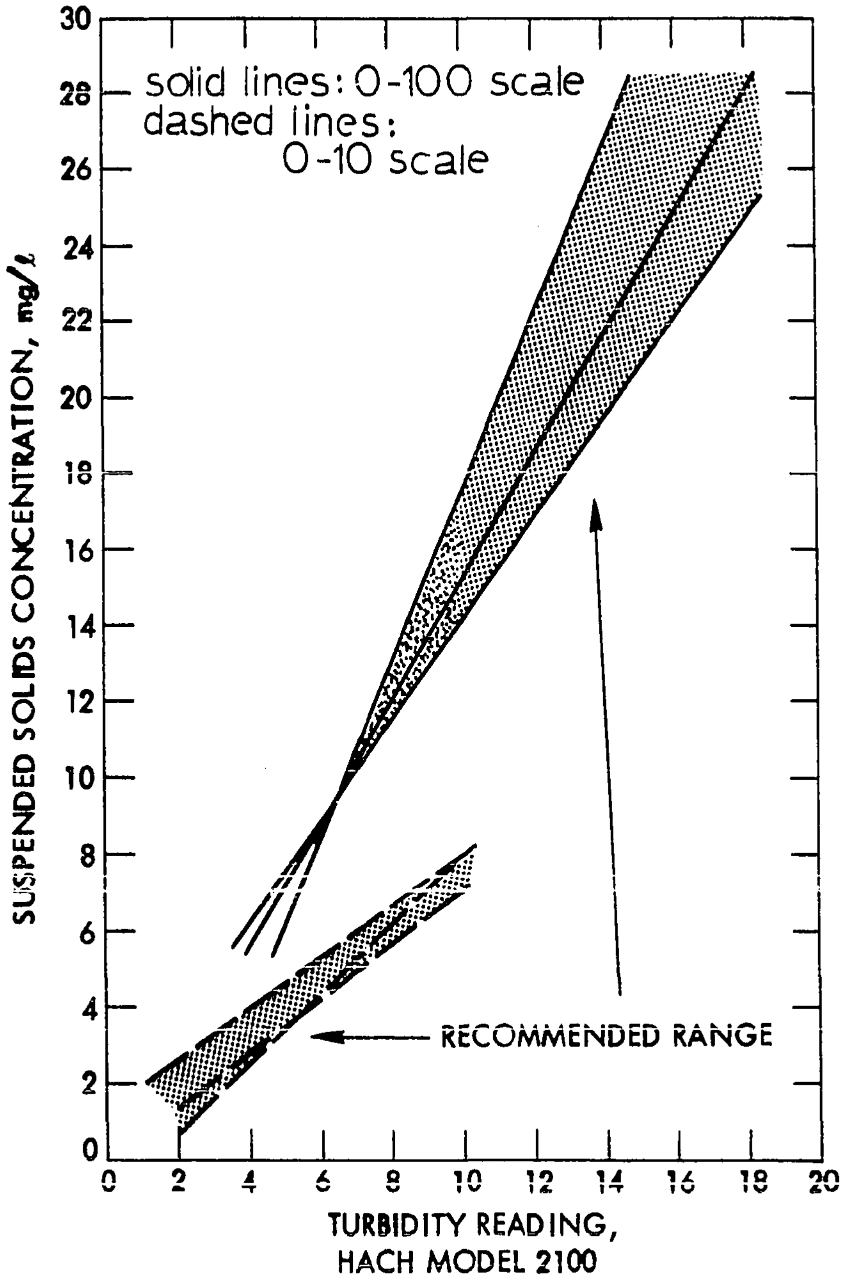

Fig. 4. Turbidity - suspended solids relationships

2. Filter run

A filter run consisted of filtering a secondary final effluent from the Ames water pollution control plant at a predetermined flow rate for a period of time sufficient to show a positive degradation of filtrate from each of the various depths of the 4-depth filter cells or until the headloss across the filter cells reached the allowable maximum head limit. The details of the operation of the pilot plant apparatus during a typical run were as follows:

a. Pre-run preparation The pump used to withdraw the wastewater from the manhole and deliver it to the storage and mixing tank was started. It took about fifty minutes for this pump with a capacity of 80 gpm to fill the tank, which has a capacity of about 4,000 gallons. The wastewater in the tank was kept uniformly mixed using a 14-in. diameter propeller turning at about 700 rpm in the center of the tank. During this time, the filter cells were backwashed with clean carbon-filtered water. After backwashing, the backwash valves were closed slowly and evenly to attain uniform minimum filter bed porosity. Special care was exerted to insure the same degree of packing in each filter cell. The required care took many repeated operations to develop.

The experimental filter runs were started immediately after the filters had been backwashed and prepared for operation. Clean water from the backwash storage tank was filtered for a few minutes at the start of each experimental run during which time entrapped air was removed from the filter system and the manometers. Actual media depths in each filter cell were recorded and manometer readings were adjusted to the zero reading.

b. Filtration When the storage and mixing tank was filled with secondary effluent, the second pump, which delivered the wastewater to the filter housing, was started. Each filter cell of the three filter sets was adjusted to a pressure of 18 psi at the gauge in the top of the filter unit by valve adjustment of the flow from the pump. It was the writer's experience that operating with this pressure in the filter housing a more uniform packing of filter media could be obtained.

The first set of influent and effluent samples were collected and the initial manometer readings were recorded 5 minutes after starting filtration. Subsequent headloss readings and wastewater samples were taken every half hour for the first two hours of the run and at one hour intervals thereafter. Headloss was recorded in inches of the fluid used in the manometer and converted later to feet of water. Due to the large number of turbidity measurements (12 filter cells in operation at the same time) which needed to be made during the early period of the run, a working schedule was developed as shown in Table 3. By following this working schedule, the complicated sampling schedule became routine.

As shown in the working schedule, at four intervals during the run both turbidity and suspended solids were determined on samples of influent and effluent. Thus, a correlation between turbidity and suspended solids could be developed. Filter runs were terminated when the fourth filter cell with a 24 in. media depth reached a 10 ft. head-loss. The length of run ranged from a few hours to sixty hours depending on the filter physical characteristics, operating conditions, and the characteristics of the waste-water.

c. Backwashing On completion of a filter run, all the influent valves were closed and all the backwash valves were opened ready for backwashing. The backwashing technique used was not designed to be optimum, but was used to provide a clean filter. Much more efficient and practical techniques can be designed. All filters were backwashed for a few minutes first to break up the surface cake and to bring the floc deposited in the lower section of the filter bed to the surface. It was necessary to open the cap of each filter cell and use a gloved hand to break up the flocs which were floating on the surface of the media. These flocs were too heavy to be lifted up to the outlet which is located on the inside of the cap. Occa-

Table 3. Wastewater filtration working schedule

9:00[a]	Start filter set #1
9:05	Take reading on filter set #1
9:10	Start filter set #2
9:15	Take reading on filter set #2
9:30	Take reading on filter set #1
9:40	Take reading on filter set #2
9:50	Start filter set #3
9:55	Take reading on filter set #3
10:00	Take reading on filter set #1
10:10	Take reading on filter set #2
10:20	Take reading on filter set #3
10:30	Take reading on filter set #1
10:40	Take reading on filter set #2
10:50	Take reading on filter set #3
11:00	Take reading on filter set #1
11:10	Take reading on filter set #2
11:20	Take reading on filter set #3
11:50	Take reading on filter set #3
12:00	Take reading on filter set #1
12:10	Take reading on filter set #2
	S.S. Test
12:50	#3
1:00	#1
1:10	#2
	S.S. Test
1:50	#3
2:00	#1
2:10	#2

[a]Run clock adjusted to start of run to show a time of 9:00 without regard to actual time.

Table 3. (Continued)

2:50	#3
3:00	#1
3:10	#2
S.S. Test	
3:50	#3
4:00	#1
4:10	#2
4:50	#3
5:00	#1
5:10	#2
S.S. Test	
5:50	#3
6:00	#1
6:10	#2

sionally air bubbles sucked through the pump agitated those flocs and they were broken into smaller pieces. Then, they were carried away by the flow, as shown in Fig. 5. This would indicate that an air wash to break up the flocs preceding the water wash is an essential requirement for effective backwashing for wastewater filters.

After breaking up deposited flocs by hand, the filter cell cap was put back and the filter was washed with water at a flow rate which expanded the bed to 50 to 100 per cent. The flow rate required varied depending on the media size. For example, a backwash rate of 45 gpm/sq. ft. was required to expand anthracite with a size of 1.84 mm, a rate of 30 gpm/sq. ft. to expand anthracite with a size of 1.09 mm, both to a 50 per cent bed expansion. During this study, it was found that a backwash rate which was sufficient to expand the bed 50 to 100 per cent for 3 to 5 minutes would clean the bed thoroughly, if deposited flocs had been air-agitated or broken up into sufficiently small particles.

Other components of the filtration system, such as the inlet line and rotameter were flushed thoroughly with clean water to remove possible deposits of flocs. Occasional additions of a small amount of Clorox during backwashing enhanced the cleaning process.

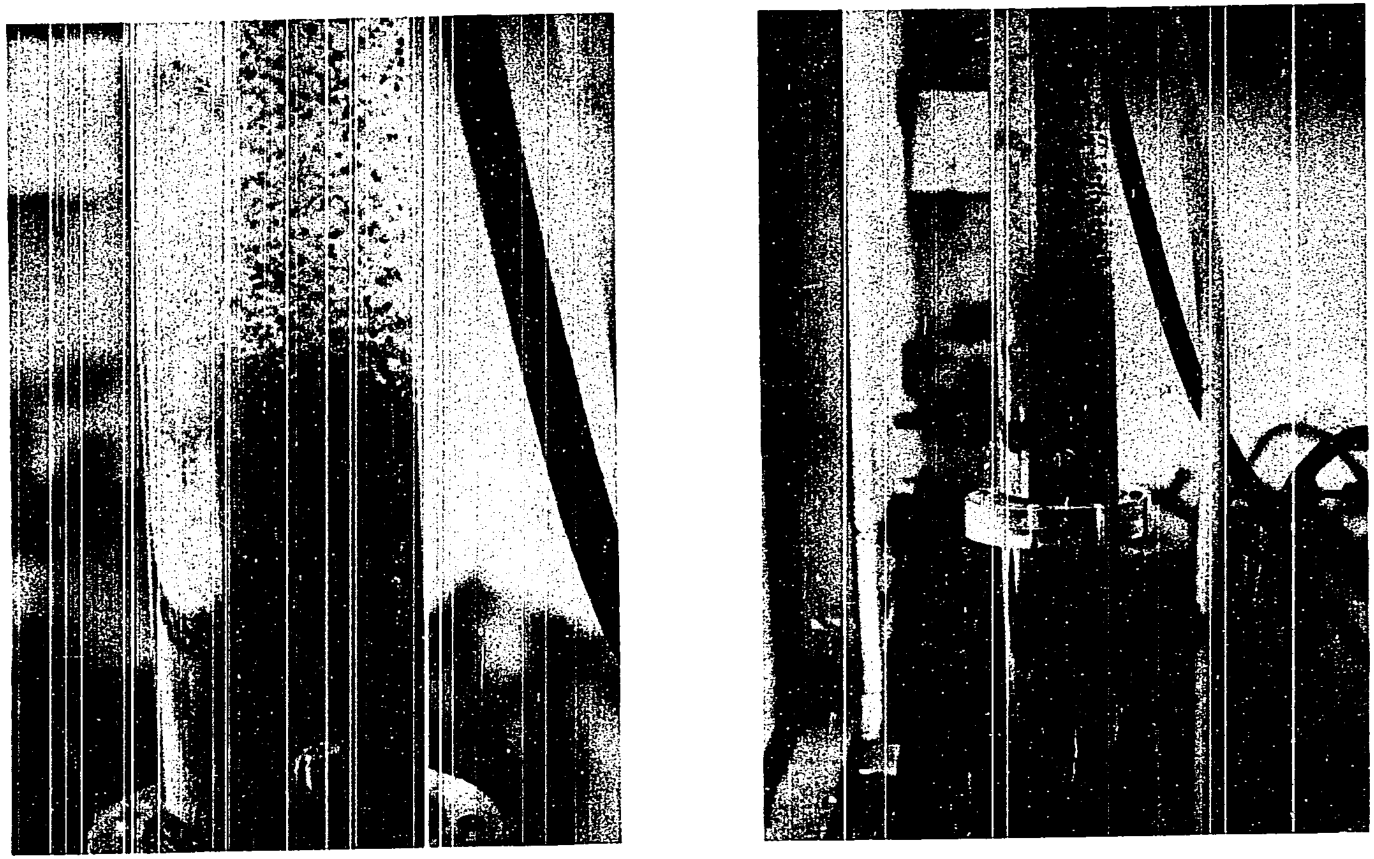

Fig. 5. Backwashing of the filter.

E. Data Analysis

In order to simplify the analysis of all the data, a computer program designated as "Analysis Program with Simplotter Subroutines" was written in Fortran language. The Analysis Program consists of three steps: feeding in the information, conducting calculations, and outputting the results.

1. Feeding in information

The input data for a filter run fed into the program are as follows:

* Identification information - number and data of run,
* Run characteristics - water temperature, flow rate, media size and distribution, and media depth of each filter cell,
* Filter performance - effluent quality and head loss for each cell measured from the time of beginning of filtration.

2. Calculations

The computer program was designed to feed in the information described in the previous section and to calculate the following:

* Filter efficiency, C/C_o, at various depths in a filter and at various times in a filter run.

* Volume of water produced per unit surface area by multiplying flow rate, gpm/sq. ft., by the time since the beginning of the run, hr.
* Accumulation of suspended solids inside the filter pores by the relationship

$$S_a\left(\frac{\text{gram}}{\text{cu. ft.}}\right) =$$

$$\frac{(C_o - C)\,(^{mg}/\ell) \times v\,(^{gpm}/\text{sq.ft.}) \times 3.785\,(^{\ell}/\text{gal.}) \times 60\,(^{min.}/\text{hr.}) \times t(\text{hr.})}{1(\text{in.}) \times \frac{1}{12}\,(^{ft.}/\text{in.})}$$

* Filter coefficient, λ, which was calculated by the relationship

$$\lambda = \frac{\ln {}^{C_1}/_{C_2}}{l_2 - l_1}$$

* Rate of headloss build-up, ft. of water column, per unit filter depth, in., and
* Solids removal rate, $\Delta C/\Delta l/C$, fractional removal rate in SS per unit depth.

3. Outputting information

The computer output included raw data as shown in Table 4 and data analysis as shown in Table 5. Table 4 shows the raw data of run F-1-II, in which F designates a run in the Phase F study (investigation of flow rate effect on dual media filter performance), 1 represents the first run in that phase and II represents filter set II. In this run, the filter media was 1.84 mm anthracite coal on top of 0.55 mm sand; the average influent suspended solids was 12.5 mg/ℓ; water temperature was 21.5°C, and the flow rate was 4 gpm/sq. ft. A typical detailed summary of removal efficiency (C/C_o), suspended solids accumulation (S_a), filter coefficient (λ), headloss rate (dh/dl), and removal rate (dc/dl/C) in each segment along the filter bed is presented in Table 5.

The results of this dissertation presented in the next chapters were obtained chiefly from analysis of the computer output described above.

Table 4. Raw data - granular filter for wastewater filtration

RUN NO. F-1-2 , DATE 7-26-71 , CO= 12.5 MG/L, WATER TEMP= 21.5 C
FLOW RATE= 4.0 GPM/SQ FT, COAL SIZE= 1.84 MM, SAND SIZE= 0.55 MM
NO CHEMICAL ADDED

		* COLUMN NO.1 * DEPTH= 1.00 IN			* COLUMN NO.2 * DEPTH= 7.75 IN			* COLUMN NO.3 * DEPTHC=12.50 IN DEPTHS= 3.00 IN			* COLUMN NO.4 * DEPTHC=12.50 IN DEPTHS=12.00 IN			VOLUME WATER PROD.
T HR	CO MG/L	C MG/L	C/CO	H FT	C MG/L	C/CO	H FT	C MG/L	C/CO	H FT	C MG/L	C/CO	H FT	Q G/SQFT
0.1	12.5	8.5	0.68	0.08	4.90	0.39	0.11	2.10	0.17	0.31	1.50	0.12	0.94	24.0
0.5	12.5	8.3	0.66	0.08	4.10	0.33	0.11	2.70	0.22	0.34	2.30	0.18	0.97	120.0
1.0	12.5	6.3	0.50	0.08	3.10	0.25	0.11	1.90	0.15	0.37	1.30	0.10	1.00	240.0
1.5	11.1	6.1	0.55	0.10	2.50	0.23	0.13	1.70	0.15	0.37	1.30	0.12	1.03	360.0
4.0	12.1	7.1	0.59	0.19	2.10	0.17	0.26	1.50	0.12	0.47	1.30	0.11	1.21	960.0
5.0	12.1	11.3	0.93	0.28	1.90	0.16	0.35	1.70	0.14	0.60	1.50	0.12	1.36	1200.0
7.0	12.9	8.1	0.63	0.48	1.90	0.15	0.62	1.30	0.10	0.85	1.10	0.09	1.68	1680.0
8.0	13.4	9.5	0.71	0.59	2.90	0.22	0.83	2.70	0.20	1.01	2.10	0.16	1.89	1920.0
9.0	12.9	10.3	0.80	0.74	3.50	0.27	0.97	2.70	0.21	1.18	2.30	0.18	2.10	2160.0
10.0	12.9	10.7	0.83	0.89	3.30	0.26	1.26	2.50	0.19	1.42	2.30	0.18	2.41	2400.0
11.0	12.9	10.1	0.78	1.14	2.90	0.22	1.39	2.70	0.21	1.68	2.30	0.18	2.68	2640.0
12.0	14.2	10.3	0.73	1.30	2.90	0.20	1.68	2.30	0.16	1.84	1.90	0.13	2.94	2880.0
22.0	14.0	11.9	0.85	3.68	1.50	0.11	4.46	1.10	0.08	4.72	0.90	0.06	6.19	5280.0
24.0	11.1	9.3	0.84	4.10	1.10	0.10	5.22	0.70	0.06	5.41	0.50	0.05	6.98	5760.0
28.0	11.7	9.5	0.81	4.99	0.90	0.08	6.50	0.70	0.06	7.19	0.50	0.04	8.45	6720.0
35.0	11.3	11.1	0.98	6.67	2.30	0.20	9.58	0.70	0.06	10.45	0.50	0.04	11.81	8400.0

Table 5. Data analysis - granular filter for wastewater filtration

RUN NO. F-1-2 , DATE 7-26-71 ,CO= 12.5 MG/L, WATER TEMP= 21.5 C
FLOW RATE= 4.0 GPM/SQ FT, COAL SIZE= 1.84 MM, SAND SIZE= 0.55 MM
NO CHEMICAL ADDED

WITHIN FILTER COLUMN FROM 0.0 IN TO 1.0 IN

T HR	CO,SS MG/L	C,SS MG/L	C/CO	SS ACCUM GM/CU FT	FILT COEF /IN	HEADLOSS FT(WATER)	HEAD RATE FT/IN	REM RATE DC/DL/C
0.1	12.5	8.5	0.68	4.36	0.386	0.08	0.08	0.320
0.5	12.5	8.3	0.66	22.24	0.409	0.08	0.08	0.336
1.0	12.5	6.3	0.50	50.58	0.685	0.08	0.08	0.496
1.5	11.1	6.1	0.55	81.10	0.599	0.10	0.10	0.450
4.0	12.1	7.1	0.59	217.36	0.533	0.19	0.19	0.413
5.0	12.1	11.3	0.93	248.97	0.068	0.28	0.28	0.066
7.0	12.9	8.1	0.63	310.02	0.465	0.48	0.48	0.372
8.0	13.4	9.5	0.71	357.44	0.344	0.59	0.59	0.291
9.0	12.9	10.3	0.80	392.86	0.225	0.74	0.74	0.202
10.0	12.9	10.7	0.83	419.03	0.187	0.89	0.89	0.171
11.0	12.9	10.1	0.78	446.28	0.245	1.14	1.14	0.217
12.0	14.2	10.3	0.73	482.80	0.321	1.30	1.30	0.275
22.0	14.0	11.9	0.85	809.82	0.163	3.68	3.68	0.150
24.0	11.1	9.3	0.84	852.33	0.177	4.10	4.10	0.162
28.0	11.7	9.5	0.81	939.54	0.208	4.99	4.99	0.188
35.0	11.3	11.1	0.98	1031.10	0.018	6.67	6.67	0.018

Table 5. (Continued)

RUN NO. F-1-2 , DATE 7-26-71 ,CO= 12.5 MG/L, WATER TEMP= 21.5 C
FLOW RATE= 4.0 GPM/SQ FT, COAL SIZE= 1.84 MM, SAND SIZE= 0.55 MM
NO CHEMICAL ADDED

T HR	CO,SS MG/L	C,SS MG/L	C/CO	SS ACCUM GM/CU FT	FILT COEF /IN	HEADLOSS FT(WATER)	HEAD RATE FT/IN	REM RATE DC/DL/C
WITHIN FILTER COLUMN FROM 1.0 IN TO 7.8 IN								
0.1	8.5	4.9	0.58	0.58	0.082	0.02	0.00	0.063
0.5	8.3	4.1	0.49	3.10	0.104	0.02	0.00	0.075
1.0	6.3	3.1	0.49	6.09	0.105	0.03	0.00	0.075
1.5	6.1	2.5	0.41	8.83	0.132	0.03	0.00	0.087
4.0	7.1	2.1	0.30	26.19	0.180	0.06	0.01	0.104
5.0	11.3	1.9	0.17	37.82	0.264	0.08	0.01	0.123
7.0	8.1	1.9	0.23	63.01	0.215	0.14	0.02	0.113
8.0	9.5	2.9	0.31	73.35	0.176	0.23	0.03	0.103
9.0	10.3	3.5	0.34	84.17	0.160	0.23	0.03	0.098
10.0	10.7	3.3	0.31	95.64	0.174	0.36	0.05	0.102
11.0	10.1	2.9	0.29	107.43	0.185	0.24	0.04	0.106
12.0	10.3	2.9	0.28	119.21	0.188	0.38	0.06	0.106
22.0	11.9	1.5	0.13	262.94	0.307	0.79	0.12	0.129
24.0	9.3	1.1	0.12	292.98	0.316	1.13	0.17	0.131
28.0	9.5	0.9	0.09	347.24	0.349	1.51	0.22	0.134
35.0	11.1	2.3	0.21	445.59	0.233	2.91	0.43	0.117

Table 5. (Continued)

RUN NO. F-1-2 , DATE 7-26-71 ,CO= 12.5 MG/L, WATER TEMP= 21.5 C
FLOW RATE= 4.0 GPM/SQ FT, COAL SIZE= 1.84 MM, SAND SIZE= 0.55 MM
NO CHEMICAL ADDED

WITHIN FILTER COLUMN FROM 7.8 IN TO15.5 IN

T HR	CO,SS MG/L	C,SS MG/L	C/CO	SS ACCUM GM/CU FT	FILT COEF /IN	HEADLOSS FT(WATER)	HEAD RATE FT/IN	REM RATE DC/DL/C
0.1	4.9	2.1	0.43	0.39	0.109	0.21	0.03	0.074
0.5	4.1	2.7	0.66	1.58	0.054	0.23	0.03	0.044
1.0	3.1	1.9	0.61	2.49	0.063	0.25	0.03	0.050
1.5	2.5	1.7	0.68	3.19	0.050	0.24	0.03	0.041
4.0	2.1	1.5	0.71	5.65	0.043	0.21	0.03	0.037
5.0	1.9	1.7	0.89	6.22	0.014	0.25	0.03	0.014
7.0	1.9	1.3	0.68	7.34	0.049	0.23	0.03	0.041
8.0	2.9	2.7	0.93	7.90	0.009	0.18	0.02	0.009
9.0	3.5	2.7	0.77	8.61	0.033	0.20	0.03	0.029
10.0	3.3	2.5	0.76	9.73	0.036	0.16	0.02	0.031
11.0	2.9	2.7	0.93	10.44	0.009	0.29	0.04	0.009
12.0	2.9	2.3	0.79	11.00	0.030	0.16	0.02	0.027
22.0	1.5	1.1	0.73	18.03	0.040	0.26	0.03	0.034
24.0	1.1	0.7	0.64	19.16	0.058	0.18	0.02	0.047
28.0	0.9	0.7	0.78	20.85	0.032	0.69	0.09	0.029
35.0	2.3	0.7	0.30	29.71	0.153	0.87	0.11	0.090

Table 5. (Continued)

RUN NO. F-1-2 , DATE 7-26-71 ,CO= 12.5 MG/L, WATER TEMP= 21.5 C
FLOW RATE= 4.0 GPM/SQ FT, COAL SIZE= 1.84 MM, SAND SIZE= 0.55 MM
NO CHEMICAL ADDED

WITHIN FILTER COLUMN FROM15.5 IN TO24.5 IN

T HR	CO,SS MG/L	C,SS MG/L	C/CO	SS ACCUM GM/CU FT	FILT COEF /IN	HEADLOSS FT(WATER)	HEAD RATE FT/IN	REM RATE DC/DL/C
0.1	2.1	1.5	0.71	0.07	0.037	0.63	0.07	0.032
0.5	2.7	2.3	0.85	0.31	0.018	0.63	0.07	0.016
1.0	1.9	1.3	0.68	0.62	0.042	0.63	0.07	0.035
1.5	1.7	1.3	0.76	0.92	0.030	0.66	0.07	0.026
4.0	1.5	1.3	0.87	1.83	0.016	0.73	0.08	0.015
5.0	1.7	1.5	0.88	2.07	0.014	0.77	0.09	0.013
7.0	1.3	1.1	0.85	2.56	0.019	0.83	0.09	0.017
8.0	2.7	2.1	0.78	3.04	0.028	0.88	0.10	0.025
9.0	2.7	2.3	0.85	3.65	0.018	0.92	0.10	0.016
10.0	2.5	2.3	0.92	4.01	0.009	1.00	0.11	0.009
11.0	2.7	2.3	0.85	4.37	0.018	1.00	0.11	0.016
12.0	2.3	1.9	0.83	4.86	0.021	1.10	0.12	0.019
22.0	1.1	0.9	0.82	8.49	0.022	1.47	0.16	0.020
24.0	0.7	0.5	0.71	8.97	0.037	1.57	0.17	0.032
28.0	0.7	0.5	0.71	9.94	0.037	1.26	0.14	0.032
35.0	0.7	0.5	0.71	11.64	0.037	1.36	0.15	0.032

V. CHRONOLOGY, IDENTIFICATION, AND PURPOSE OF RUNS

In designing the operation of the three sets of 4-filter cells to collect data necessary to meet the objectives of this thesis, it was necessary to be able to identify the various runs made. A total of over 31 separate runs were made in six phases, each of which was designed to accomplish a specific purpose:

Phase A. Effect of media size on filtration of wastewater through a single media sand filter.

Phase B. Effect of flow rate on filtration of wastewater through a single media sand filter.

Phase C. Comparison of operating characteristics of single media versus dual media filters using wastewater.

Phase D. Effect of size of anthracite and sand on filtration of wastewater through a dual media filter.

Phase E. Effect of media size on filtration of wastewater through a single media anthracite filter.

Phase F. Effect of flow rate on filtration of wastewater through dual media of "selected size."

In each test phase, from 2 to 11 separate runs were made. Each individual run was designated with a three unit designation, for example, C-4-III. In the designation, the letter refers to the phase described above, the first number refers to the number of the run made in that phase, and the second number refers to the filter set whose data (from four filters of different depths) are presented as the results from a single filter. Thus, the designation C-4-III means the results from filter set III in the fourth run of Phase C.

In order to facilitate the orderly recording of essential data, Table 6 was prepared to give a summary of the general conditions existing in each phase of the pilot-plant studies.

A. Phase A: Study of Effect of Media Size in a Single media Sand Filter

Three filter sets (four filter cells in each filter set containing media depths of approximately 1, 5, 9, and 13 inches, as shown in Fig. 3a, 3b and 3c) were used in each filter run. Media sizes in each set are listed in Table 6. All filter sets were delivered the same wastewater and were operated at the same flow rate.

Fig. 6 shows the filter effluent quality and head-losses from various depths of a filter bed versus time from the beginning of the run. The typical results shown are from a uniform sand filter of 0.65 mm operated at a flow rate of 2 gpm/sq. ft. with an influent suspended solids of 15.3 mg/ℓ.

It is evident that most of the removal occurred in the upper layers of the filter media, resulting in the head-losses occurring there. The effluent from the top 1 inch contained about 6 mg/ℓ of SS. Approximately 61 per cent of the influent suspended solids were removed in the top inch

Table 6. Summary of six phases of pilot plant studies

Study phase	Type of media	Media size mm	Media depth in.	Variable between sets	Flow rate gpm/sq.ft.	Influent SS mg/ℓ	No. of runs
A	Sand	I 0.92 II 0.65 III 0.55	1, 5 9, 13- 20	media size	variable between runs (2,4,6)	variable between runs (11~33)	11
B	Sand	I 0.92 II 0.92 III 0.92	1, 5, 9, 13- 20	flow rate	I 2 II 4 III 6	variable between runs (24~34)	5
C	Sand Anthracite Sand	I 1.09 II 1.84 0.77 III 1.09 0.77	1, 5, 9~15, 24	media size	variable between runs (2,4,6)	variable between runs (26~38)	5
D	Anthracite sand	I 1.84 0.77 II 1.84 0.65 III 1.84 0.55	1, 5, 15, 24	media size	variable between runs (2,4,6)	variable between runs (12~22)	5

Table 6. (Continued)

Study phase	Type of media		Media size mm	Media depth in.	Variable between sets	Flow rate gpm/sq. ft.	Influent SS mg/ℓ	No. of runs
E	Anthracite	I	1.84	1, 8,	media size	variable between runs (2,4,6)	variable between runs (7~22)	3
		II	1.30	15,				
		III	1.09	24				
F	Anthracite - Sand	I	1.84 0.55	1, 8,	flow rate	I 6	variable between runs (12.5~12.9)	2
		II	1.84 0.55	15,		II 4		
		III	1.84 0.55	24		III 2		

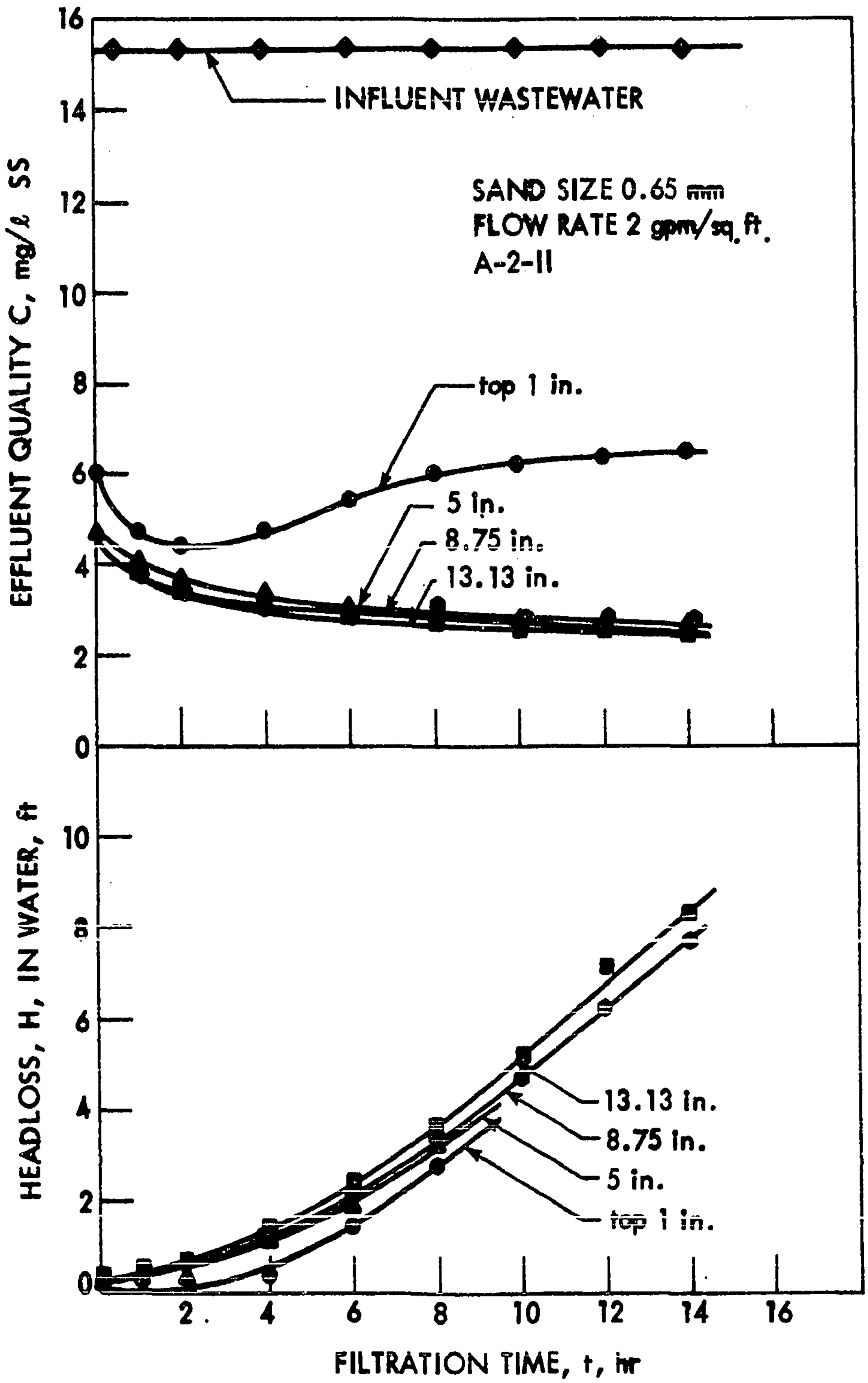

Fig. 6. Effluent quality and headloss vs. time for a single sand filter

of the filter. Providing 4 additional inches of filter media results in only a 20 per cent greater suspended solids removal. Filter media beyond a depth of 5 inches did not contribute further suspended solids removal, which made the lower portion of the filter bed relatively unutilized.

The existence of surface removal in this single media sand filter also can be revealed from the pattern of headloss build-up. As shown in Fig. 6, nearly all of the headloss occurred in the top 1 inch layer and the exponential shape of the headloss versus filtration time curve indicates that surface removal predominated in this single media sand filter.

In addition to the study of solids removal behavior and the pattern of headloss, the purpose of this phase of the study was to investigate the effect of sand size on filter performance. The pilot plant was arranged so that all variables in a single run except the sand size in the system were held constant. This tested the sensitivity of the filtration system to the one variable - sand size. It was assumed that the differences in removal efficiency and rate of headloss build-up, if any, were contributed by the difference in sand size. Flow rates studied in successive runs were 2, 4 and 6 gpm/sq. ft. A total of 11 runs were made during phase A. Fig. 7 shows the suspended solids in the influent and effluent and the headlosses observed in a

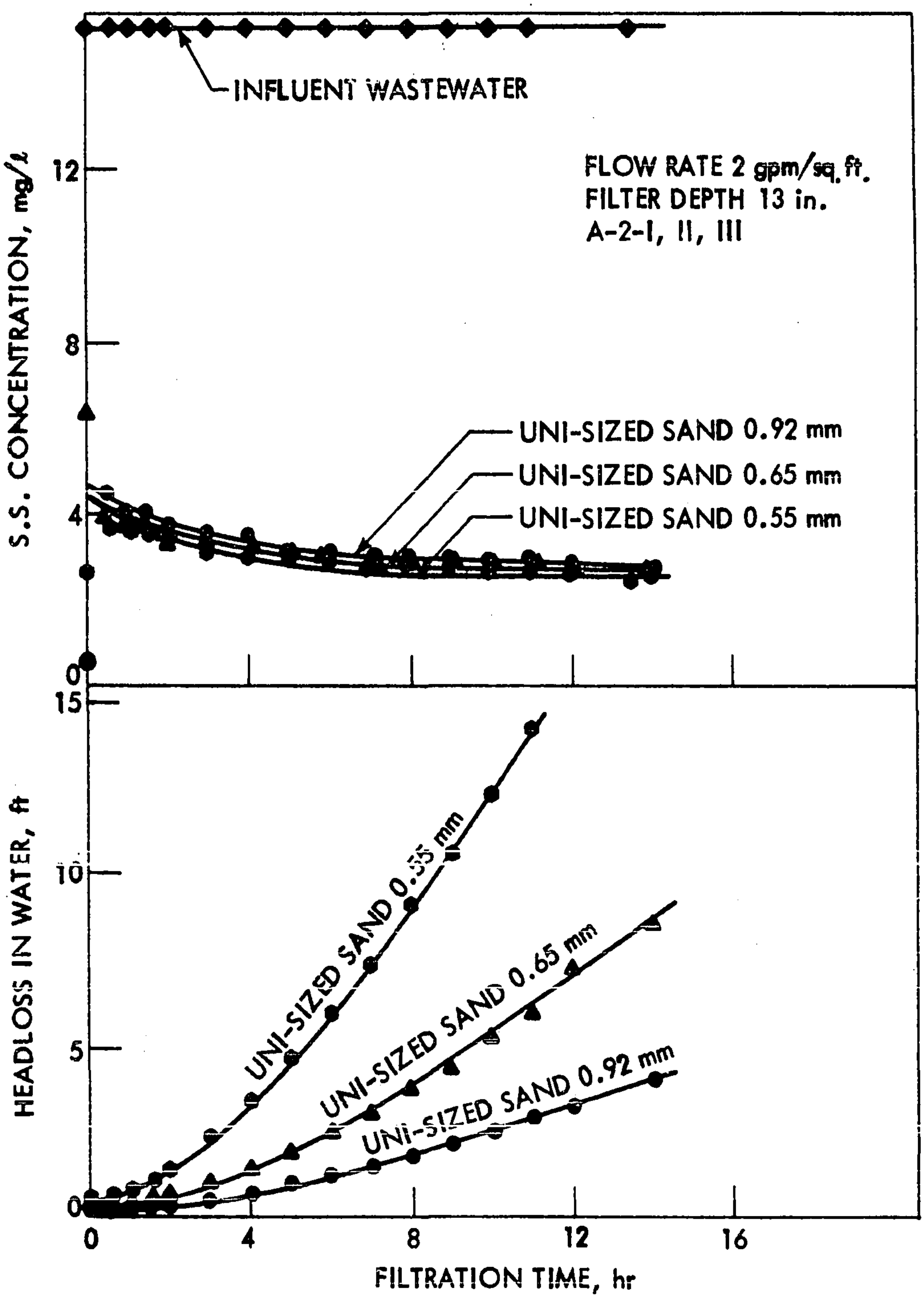

Fig. 7. Effluent quality and headloss vs. filtration time at various sand sizes

typical run made at a flow rate of 2 gpm/sq. ft. The following typical observations were made during the study of sand size effect:

(1) The three filter sets gave approximately the same effluent quality, with the finer media tending to give the better quality.

(2) The headlosses through the 0.55 and 0.65 mm sand increased so rapidly as to preclude the use of these media sizes in a single media sand filter for wastewater filtration. The headloss in the 0.92 mm sand increased rapidly, but not nearly so fast as for the other media. Therefore, it was evident that further tests should be made with 0.92 mm or a larger size.

A summary of results of phase A runs is included in Table 7.

B. Phase B: Study of Flow Rate Effect on Filter Performance

The effect of flow rate on filter performance using the 0.92 mm sand size which gave promise of providing desirable operating characteristics was studied in this phase. The only variable to be investigated during this phase was flow rate effect on filter performance. Filter sets I, II and III were operated at 2, 4 and 6 gpm/sq. ft. respectively.

Table 7. Summary of results of phase A runs

Run no.	C_o, SS mg/ℓ	Rate gpm/sq.ft.	Filtrate C*, SS mg/ℓ media size, mm 0.92	0.65	0.55	Headloss ft. water	Run length* to given headloss, hr. media size, mm 0.92	0.65	0.55
1				(trial run)					
2	15.3	2	2.9	2.8	2.7	10	20	15.5	8.5
3	20.2	4	4.2	3.9	3.8	10	5	2.5	3
4	14.0	6	3.0	3.1	2.9	10	4.5	2	3
5	11.3	4	2.9	2.6	2.5	10	6.8	4.8	3.5
6	32.5	2	6.5	6.5	6.5	10	8	3.5	2.3
7	27.2	4			(air trapped in influent)				
8	32.5	6	-	15.3	14.6	10	-	1	0.8
9	21.9	4	-	5.8	5.4	10	-	2	1.8
10	28.5	6	13.8	13.5	12.7	10	2	0.5	0.5
11	25.9	2	10.5	10	7	10	5.5	1.8	2

* Representative filtrate quality from 4th filter cell (13~20 in. depth) of each filter set.

A total of five runs, summarized in Table 8, were made during phase B of the study.

Fig. 8 shows the results of one typical run when the influent suspended solids averaged 34 mg/ℓ The results of Fig. 8 and Table 8 led to the following conclusions:

(1) The three filters operated at different rates gave approximately the same effluent quality, with the flow rate of 2 gpm/sq. ft. providing a slightly better quality. Flow rate did not significantly affect effluent quality.

(2) The headloss increase is least at the flow rate of 2 gpm/sq. ft. This might be because at a low flow rate less wastewater was filtered during the same amount of time than would have been filtered at a higher flow rate, thus resulting in a lower solids accumulation within the filter pores.

(3) Operation with pressure losses as great as 24 ft. of water did not impair filtrate quality.

Fig. 9 shows that the per cent of SS removal is lower in the top portion of the bed at higher flow rates. The higher rate causes deeper penetration of suspended solids and a higher per cent of removal in the lower portion of the filter. Apparently, the higher shear forces due to the high flow-through velocity at a high flow rate reduces the removal in the upper portion of the bed.

Table 8. Summary of results of phase B runs

Run no.	C_o, SS mg/ℓ	Media size mm	Filtrate C*, SS mg/ℓ Flow rate, gpm/sq.ft.			Headloss ft. water	Run length* to given headloss, hr. Flow rate, gpm/sq.ft.		
			2	4	6		2	4	6
1**		0.92	(2 gpm/sq. ft. for all sets)						
2	33.8	0.92	5	10.5	12	10	6	2.5	1.2
3	24.3	0.92	3.5	4.3	4.5	10	13	7.5	5
4**		0.92	(4 gpm/sq. ft. for all sets)						
5	33.6	0.92	3.5	4.5	7	10	9	5.5	3

* Representative filtrate quality from 4th filter cell (20 in. depth) of each filter set.

** Using carbon filtrate as influent to the filter.

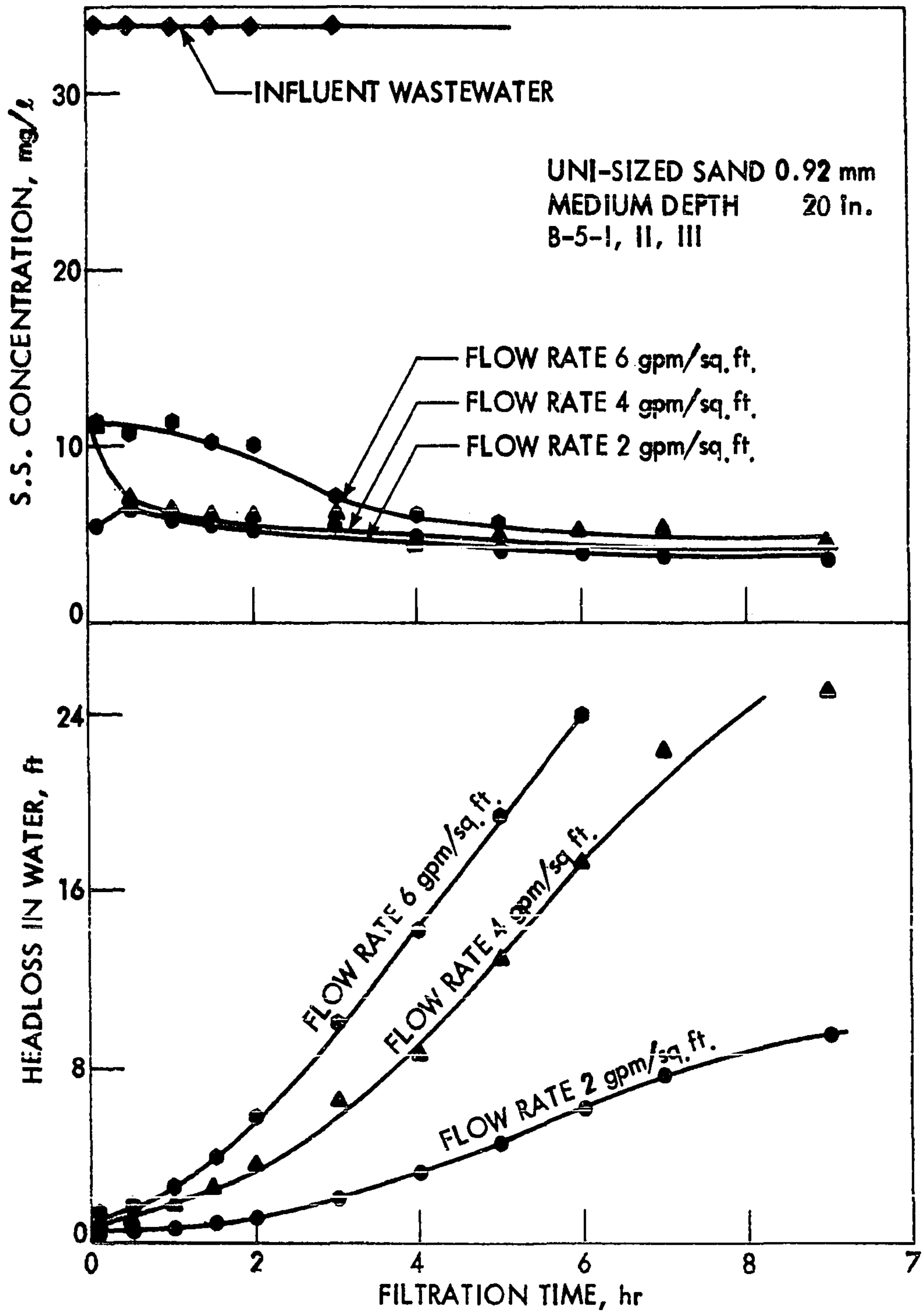

Fig. 8. Effluent quality and headloss vs. filtration time at various flow rates

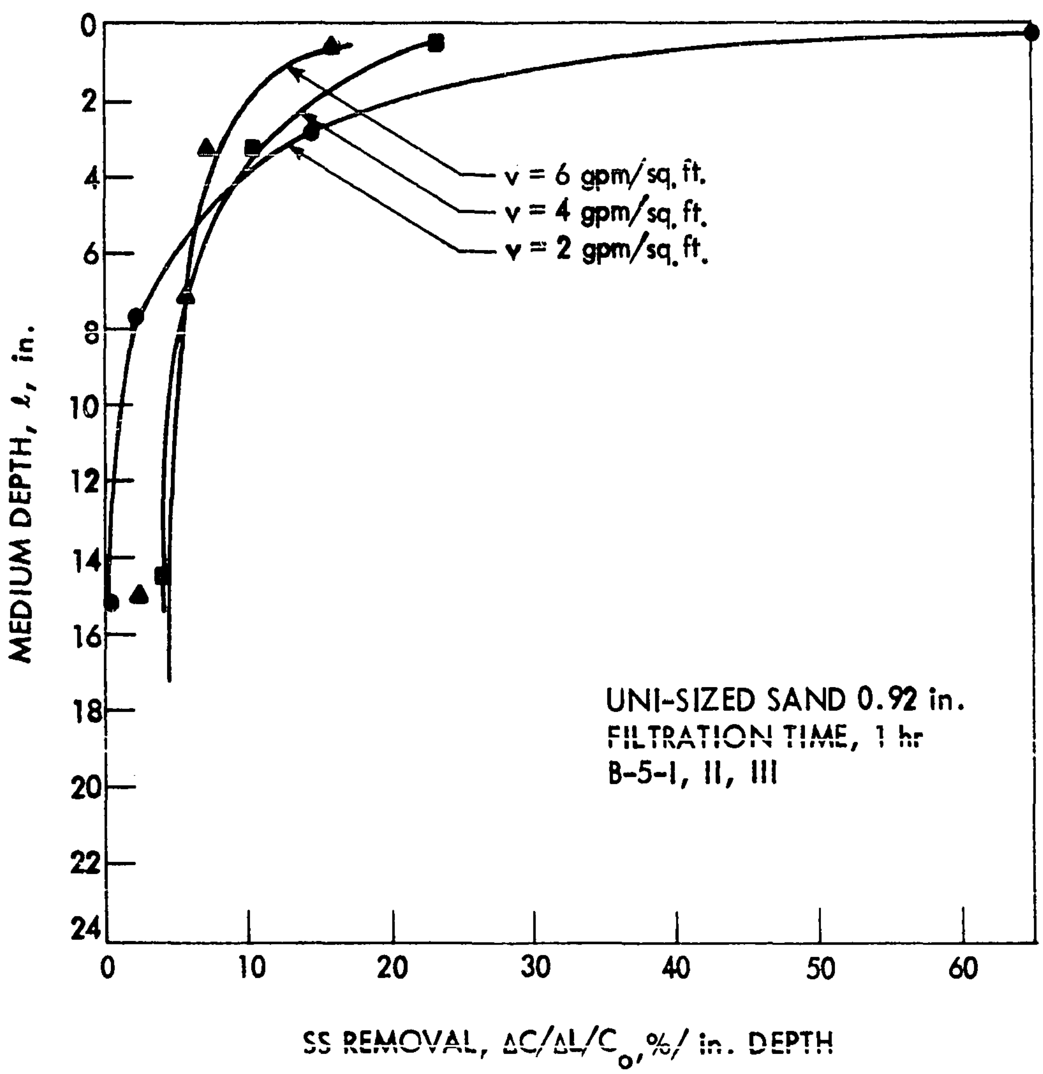

Fig. 9. Distribution of per cent of SS removal per inch of depth in uniform sand filter at various flow rates

If the maximum allowable headloss was fixed at 10 ft., the run lengths for filters with flow rates of 2, 4 and 6 gpm/sq. ft. were, respectively, 9, 4.5 and 3 hours. However, the effluent qualities at the end of the run (dictated by the headloss limit) were still within the period of "improving" effluent quality, in other words the suspended solids in the effluent were still decreasing. This indicates that with this wastewater the removal capacity of the filters was not utilized fully when the runs were terminated based on a maximum headloss as high as 24 ft. of water. This was far from an optimal filter design.

A more efficient filter design, in which the filter removal capacity would be utilized completely at the same time the maximum allowable headloss was reached, can be obtained by better distribution of deposits throughout the whole filter bed. Better distribution of deposits would occur if the coarsest medium filtered the most concentrated floc suspension and the finest medium filtered a suspension with a greatly reduced floc concentration.

C. Phase C: Comparison of Single and Dual Media Filters

Because most rapid filters contain sand of a non-uniform size, normal backwashing results in a size-graded filter medium. This stratification results in removal of the bulk of the suspended matter in the upper layers of the filter with a consequent inefficient use of the total depth of the bed. A modified distribution of filter media through the use of dual media beds of anthracite and sand, in which the coarse anthracite is exposed first to the influent flow, has been advanced to overcome this problem.

The purpose of this phase of the study was to compare the performance of single and dual media filters, with respect to effluent quality and rate of headloss development. The filters were operated using media size and flow rates observed in phases A and B which gave promise of giving desirable operational characteristics. The filter physical characteristics were:

Filter set I - uniform 1.09 mm sand

Filter set II - uniform 1.84 mm anthracite coal on top of uniform 0.77 mm sand

Filter set III - uniform 1.09 mm anthracite coal on top of uniform 0.77 mm sand

A total of 5 runs were studied at various flow rates, as summarized in Table 9.

Table 9. Summary of results of phase C runs

Run no.	C_o, SS mg/ℓ	Rate gpm/ sq.ft.	Filtrate C*, SS mg/ℓ media sizes, mm 1.09	1.84 0.77	1.09 0.77	Headloss ft. water	Run length* to given headloss, hr. media sizes, mm 1.09	1.84 0.77	1.09 0.77
1	33.5	4	5.5	5.5	5	10	1.5	13.5	5.4
2	33.8	2	3.5	3.5	3.5	10	9	24	12.5
3	46.6	4	7.8	6.7	5.5	10	4.4	10.5	7
4	26.0	6	2.9	2.2	2.5	10	5.5	14	7
5	38.3	6	5	5	5	10	5	11.5	9

* Representative filtrate quality from 4th filter cell (24 in. depth) of each filter set.

The comparison of filter performance was based on three criteria: 1) filter effluent quality; 2) headloss build-up through the filter; and 3) net amount of water produced in a fixed period of operation.

1. <u>Filter effluent quality</u>

Fig. 10 shows filter effluent quality versus time from the beginning of the run in a typical run. The effluent qualities from all three filters at a 24 inch media depth are essentially identical with slightly better filtrate from dual media filters. Laughlin and Duvall (44) and Oeben, *et al.* (67) found similar results.

The removal efficiency (C/C_o) at various depths in the filter bed has been an interesting subject to many researchers. Fig. 11 shows the suspended solids removal through the depth of the filter at 1, 3 and 5 hours from the beginning of the run. The removal efficiencies at a depth of 24 inches are close for all filters - single and dual media filters. However, a remarkable difference in removal efficiency is indicated in the upper layers. Layers of anthracite media show a superior removal efficiency to that of sand media. For instance, after 5 hours of filtration the removal efficiency of a filter with 1.09 mm sand media is 46 per cent, but the removal efficiency of 1.09 mm anthracite media is 65 per cent, both at the same filter depth of 10 inches from from the surface of the filter media. An interesting

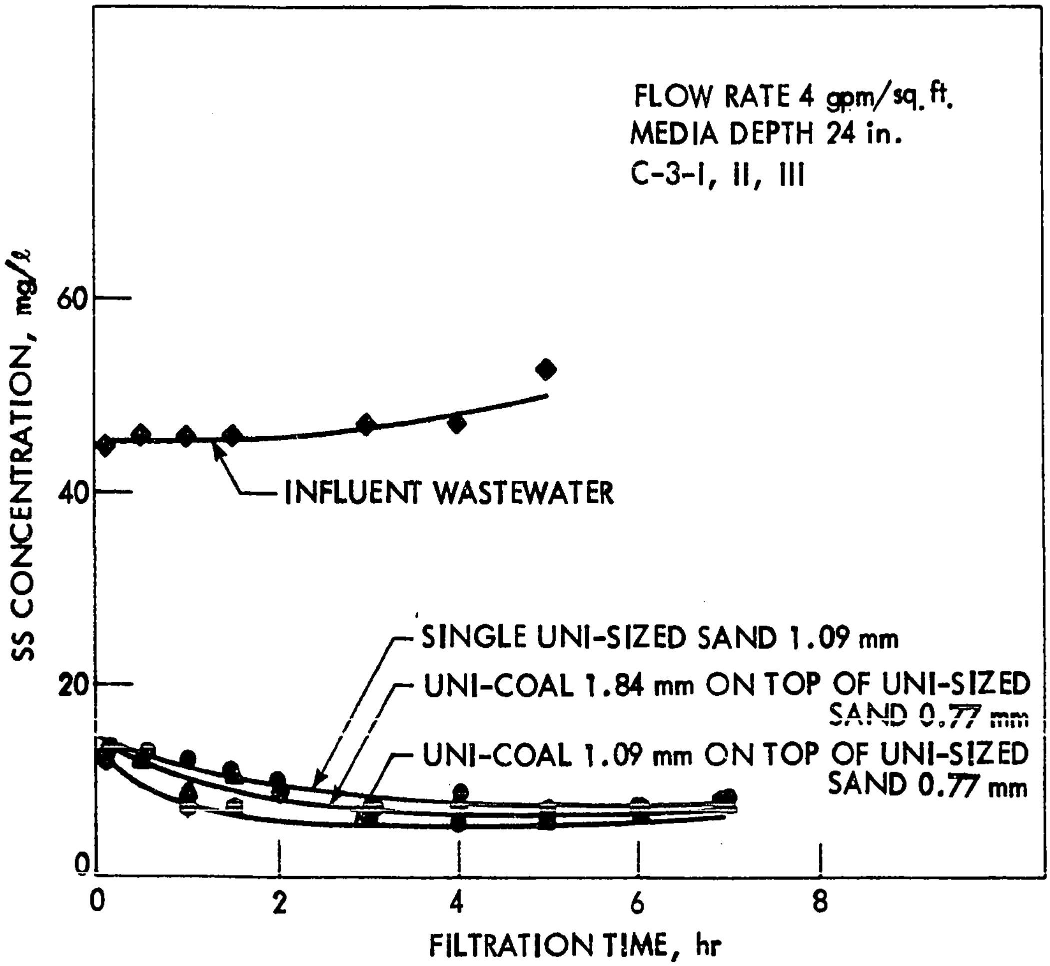

Fig. 10. Comparison of effluent quality of single and dual media filters

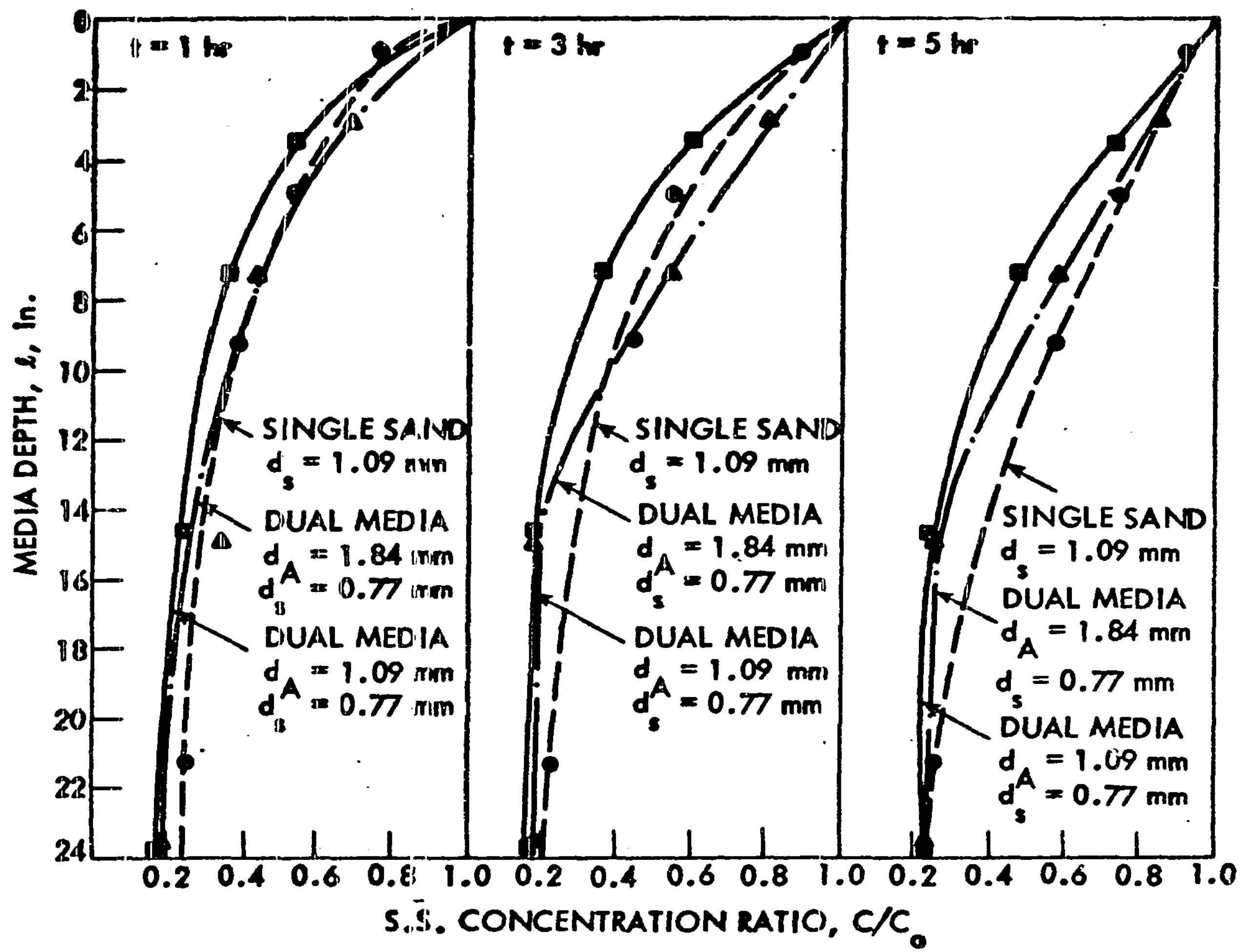

Fig. 11. Comparison of removal efficiencies of single and dual media filters at flow rate of 4 gpm/sq. ft.

observation reveals that the same media size but different grain shape and porosity (1.09 mm sand versus 1.09 mm anthracite) results in a different removal efficiency.

2. Headloss

Fig. 12 shows a typical pattern of headloss across the filter bed as the filter run progressed. The headloss developed in the single media filter is significantly higher than those in the dual media filters. As indicated in the figure, headloss reaches 10 feet of water after a 4.5 hour run length in the single-media sand filter; while it takes 7 hours for a filter with the same size anthracite and 10.7 hours for a filter with anthracite of 1.84 mm to reach a headloss of 10 feet. This slower build-up in headloss due to dual media design results in a longer filter run. In Fig. 13, the details of headloss occurrence with filter depth are demonstrated to show the differences in headloss build-up due to the different design of the filter bed.

The headlosses in the dual media filters were found to be generally only 50 per cent and 25 per cent respectively of that developed in a single media filter at the same flow rate.

If the above headloss characteristics are achieved for an equal filtrate quality, an inference can be drawn that the dual media filter is a more efficient filter. In fact, as indicated in Figs. 10 and 11, the effluent qualities for

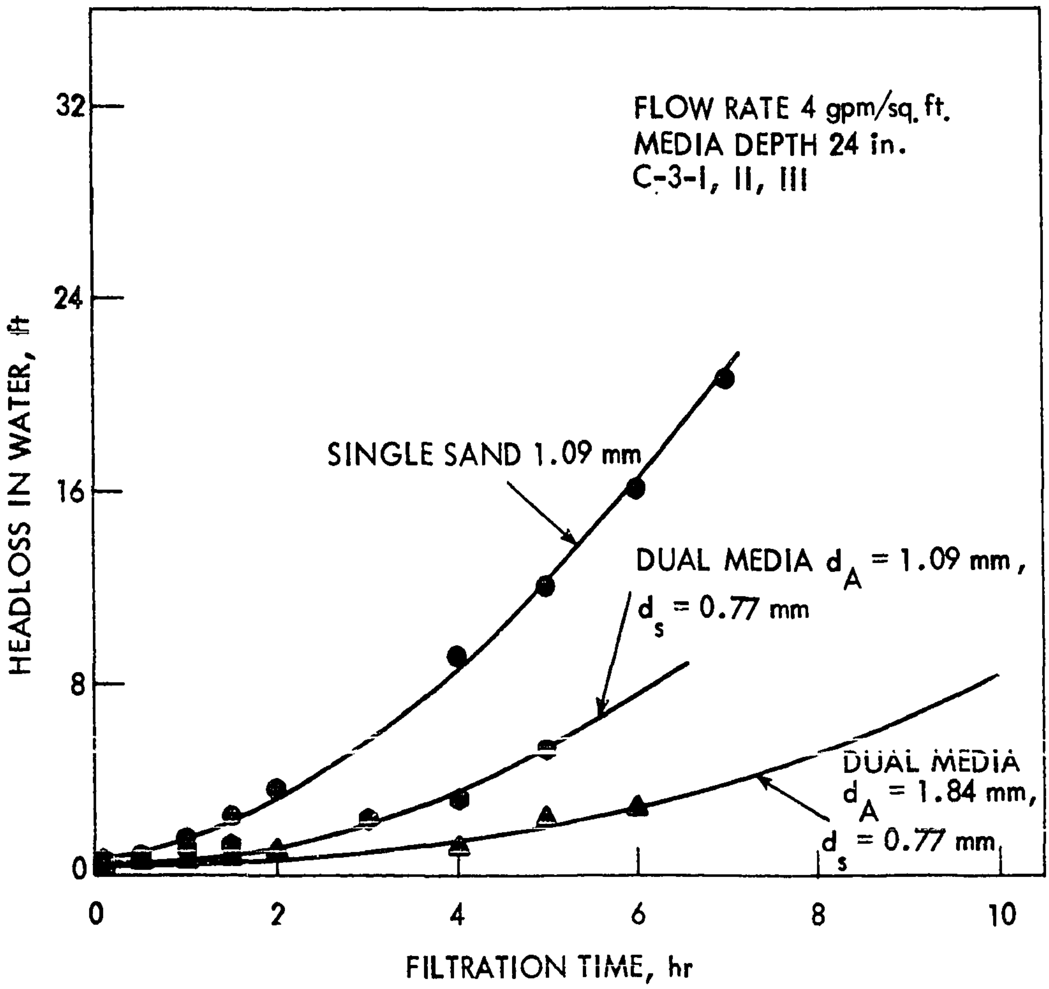

Fig. 12. Comparison of headlosses of single and dual media filters

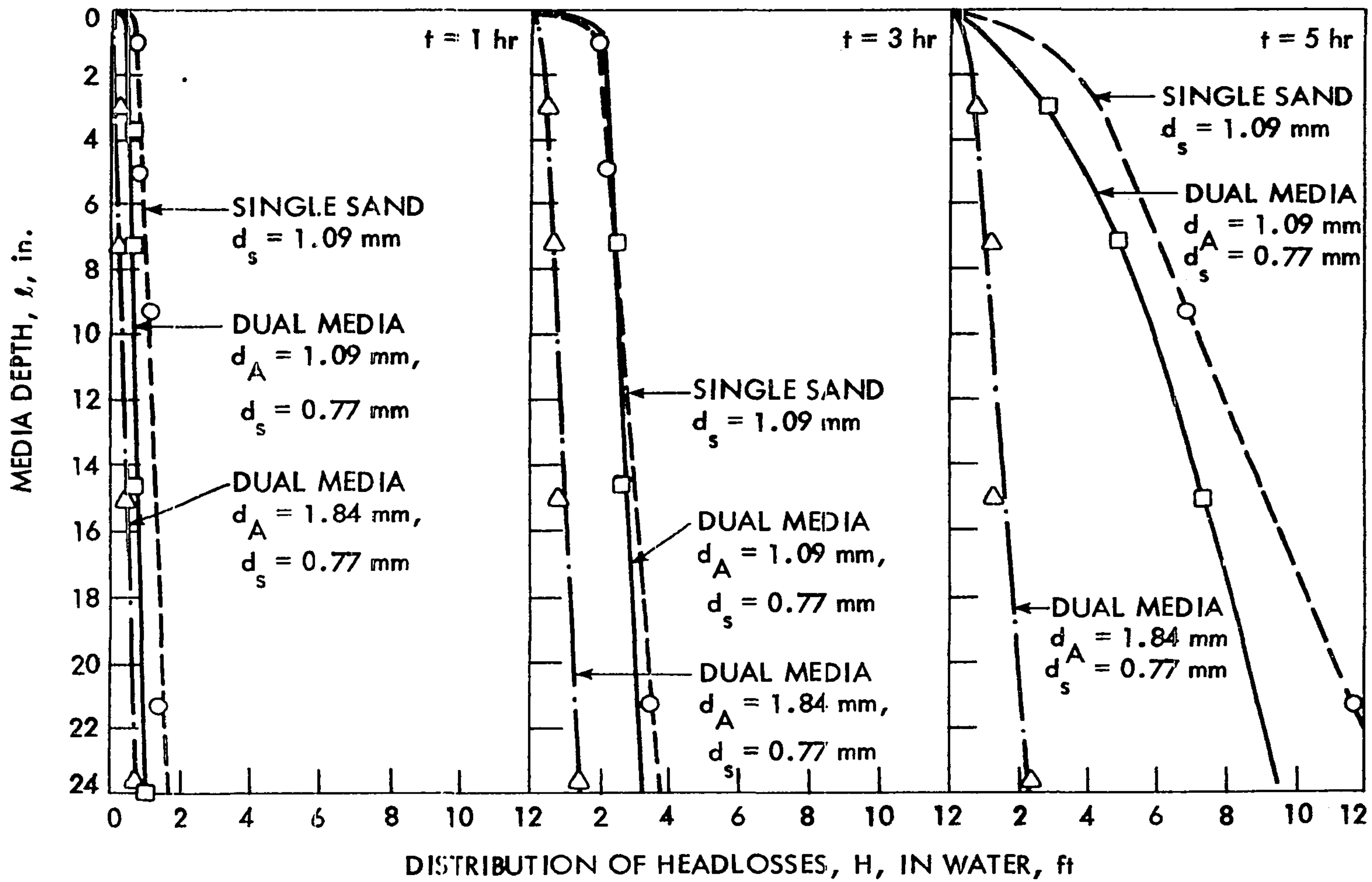

Fig. 13. Comparison of distribution of headlosses of single and dual media filters at various filtration times. C-3-I, II, III

the dual media filter were indeed superior to that of the sand, affirming the higher efficiency.

Filters with a slower headloss build-up result in a longer run length which, in turn, result in a larger amount of water production in a given period of time.

3. Quantity of water production

The quantities of water produced from the dual media filters and the single-media sand filter are compared based on their performance under an optimum operating condition, which occurs when both the headloss and effluent quality reach their respective limiting values at the same time.

The example for the determination of operational optimum design for single media sand filter and dual media anthracite-sand filter was made for Run No. C-3, in which filters were operated at a flow rate of 4 gpm/sq. ft., when influent suspended solids concentrations were 45.5 mg/ℓ. Figs. 14a, b, and c show both the filtrate quality and headlosses occurring within the filter bed and throughout the time of the filter run. As shown in these figures, the filtrate quality falls after one or two hours, remains steady for a considerable length of time, and then starts to deteriorate. Headlosses increase with depth within the filter bed and as filtration time increases. The operational optimum can be determined by drawing from Figs. 14a, b, and c the criteria for the limiting filtrate quality (C_c) and for the

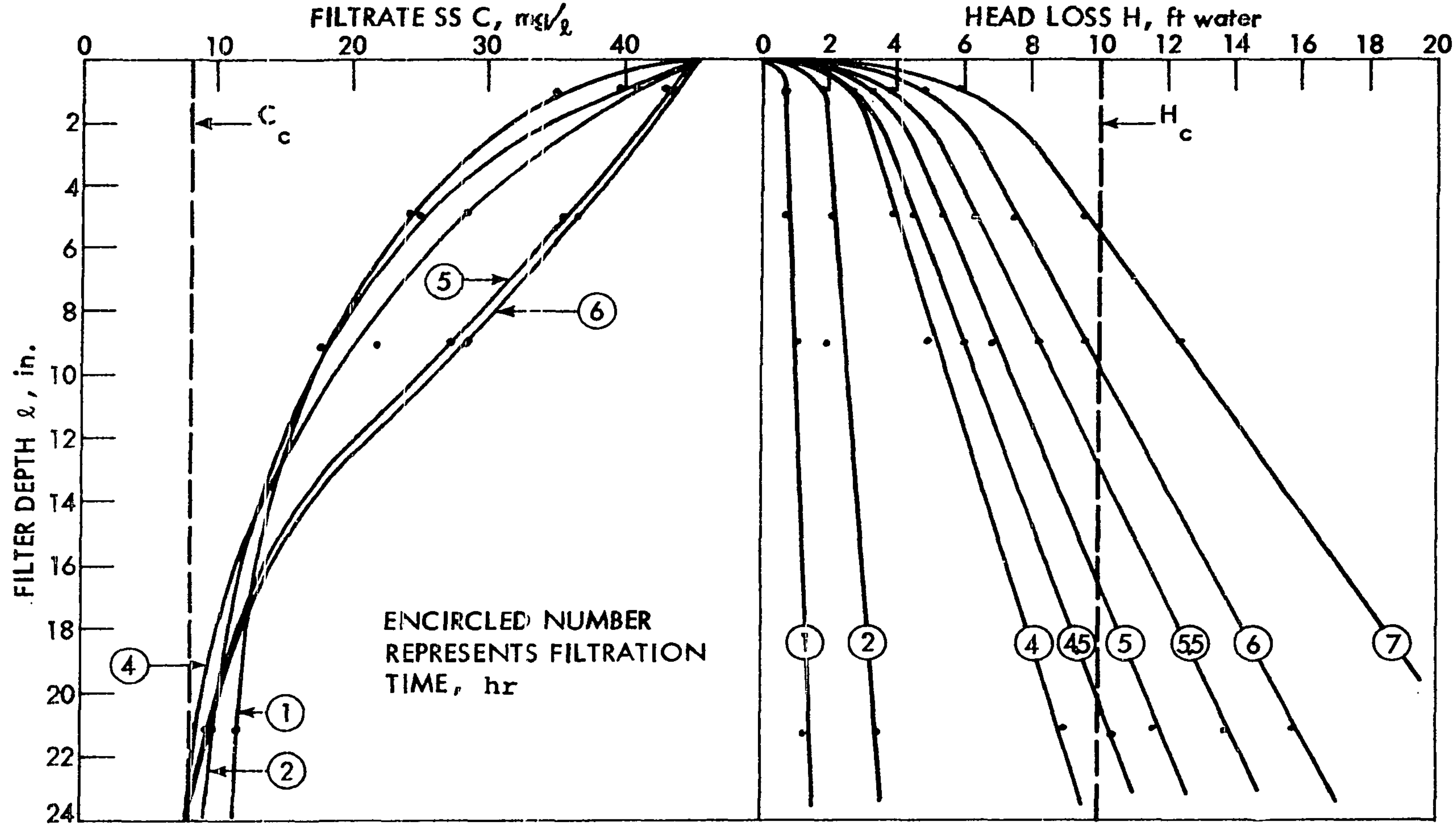

Fig. 14a. Curves of filtrate and headloss varying with depth in the filter bed and with time of filter run. Flow rate: 4 gpm/sq. ft., C-3-I

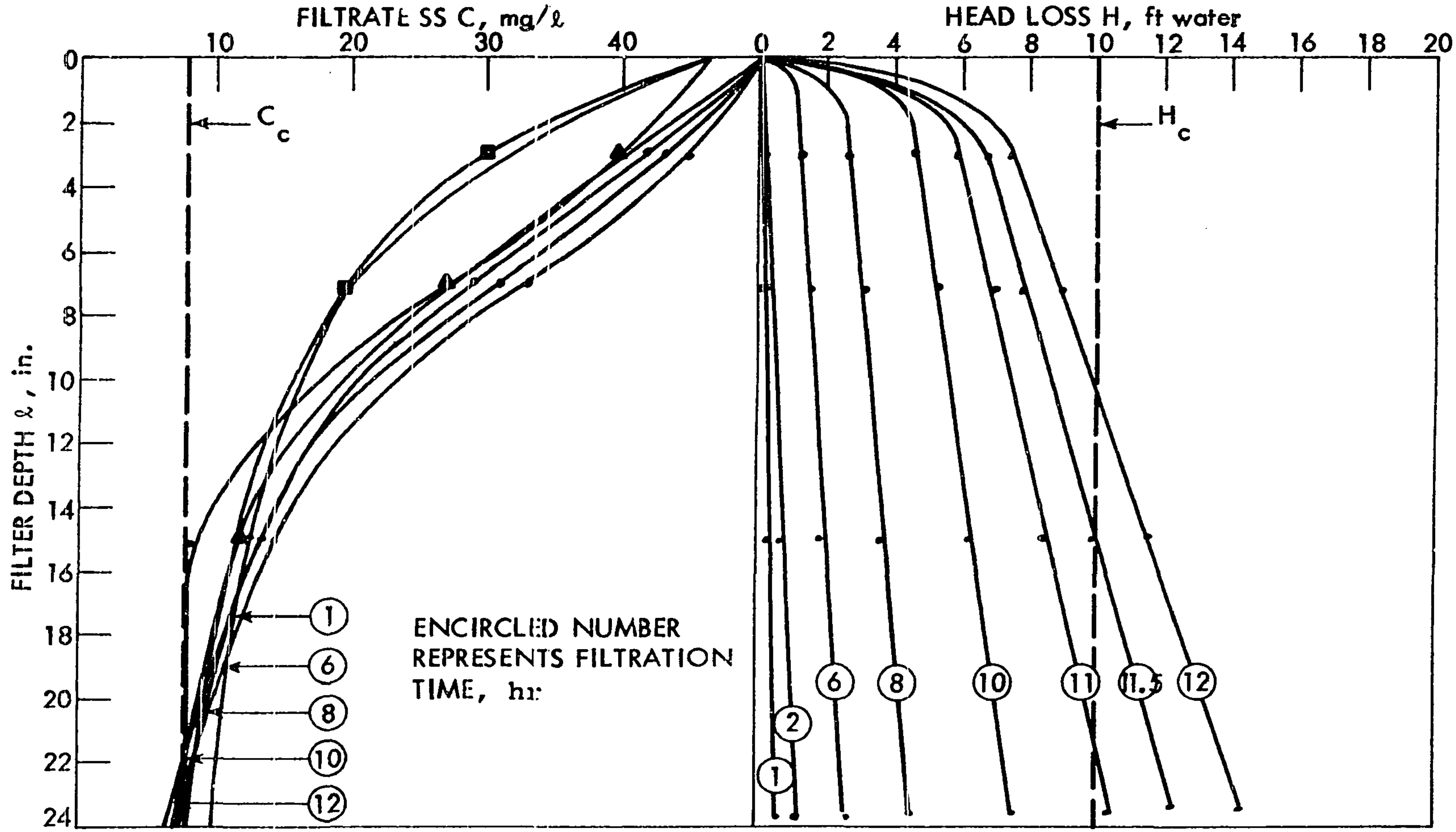

Fig. 14b. Curves of filtrate and headloss varying with depth in the filter bed and with time of filter run. Flow rate: 4 gpm/sq. ft., C-3-II

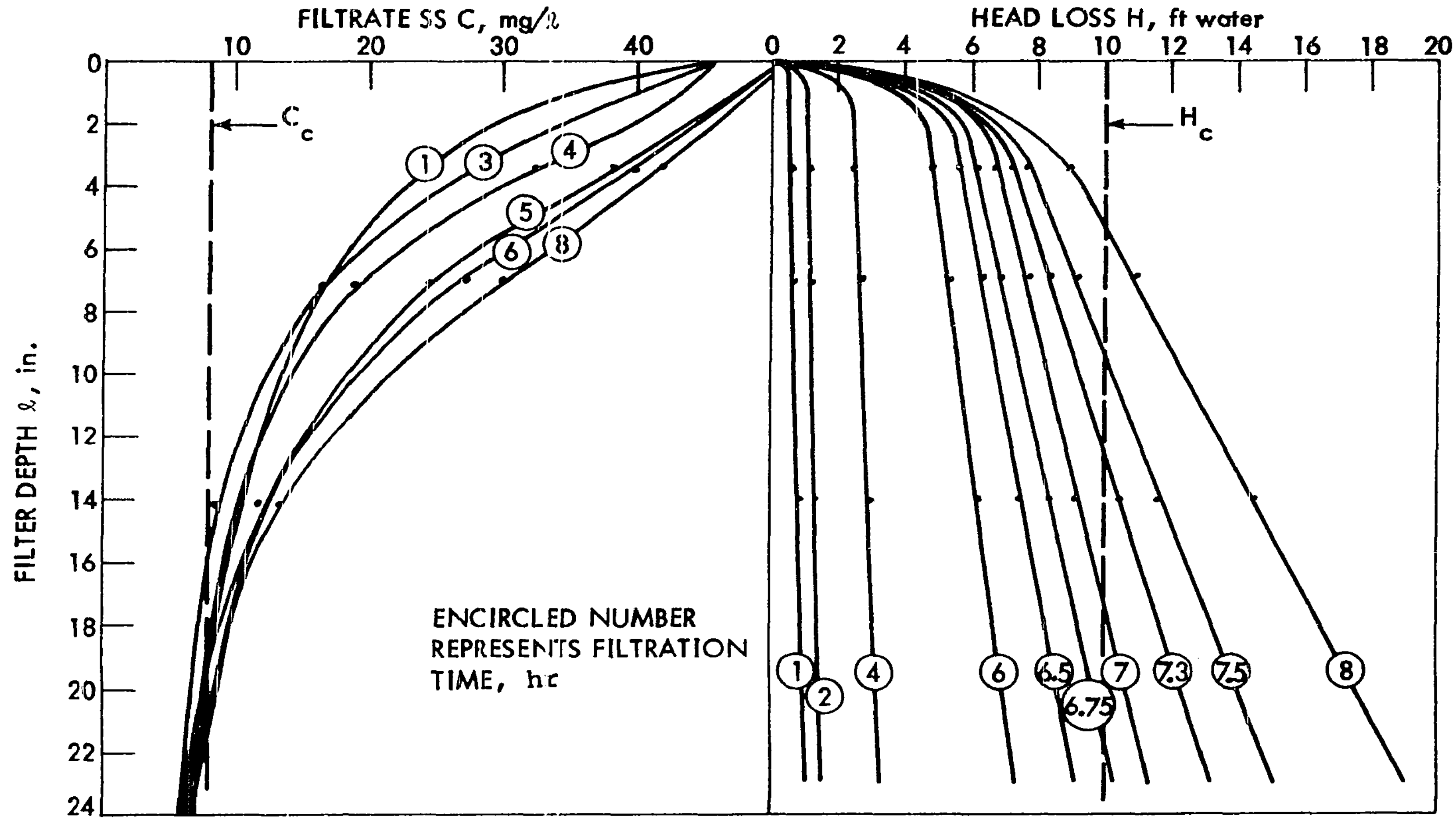

Fig. 14c. Curves of filtrate and headloss varying with depth in the filter bed and with time of filter run. Flow rate: 4 gpm/sq. ft., C-3-III

limiting headloss (H_c), as shown in the dashed lines. In this example, the limiting filtrate quality was set as 8 mg/ℓ SS, and limiting headloss was assumed to be 10 ft. water, since a terminal headloss of 10 ft. water was found to be in the range of optimal headloss in an earlier study (27). These dashed lines intercept the filtrate quality and run time where the limiting values are reached. The values of time with corresponding values for depth can be plotted as curves, as shown in Fig. 15, which is based on the data from Figs. 14a, b, and c.

Point A in Fig. 15 represents the interception of H_c and C_c curves for uni-sized 1.09 mm sand (from Fig. 14a); point B represents the interception of H_c and C_c curves for 12 in. of uni-sized 1.84 mm anthracite on top of uni-sized 0.77 mm sand (from Fig. 14b); and point C represents the interception of H_c and C_c curves for 12 in. of uni-sized 1.09 mm anthracite on top of uni-sized 0.77 mm sand (from Fig. 14c). Points A, B, and C in Fig. 15 define the optimum filter run lengths and the optimum media depths (including 12 in. anthracite on top) for the three filters under comparison, at a flow rate of 4 gpm/sq. ft. and influent suspended solids concentrations of 45.5 mg/ℓ.

The results indicated from Fig. 15 show that the optimum filter depths for a single media 1.09 mm sand filter, a 1.84 mm anthracite - 0.77 mm sand dual media filter and a

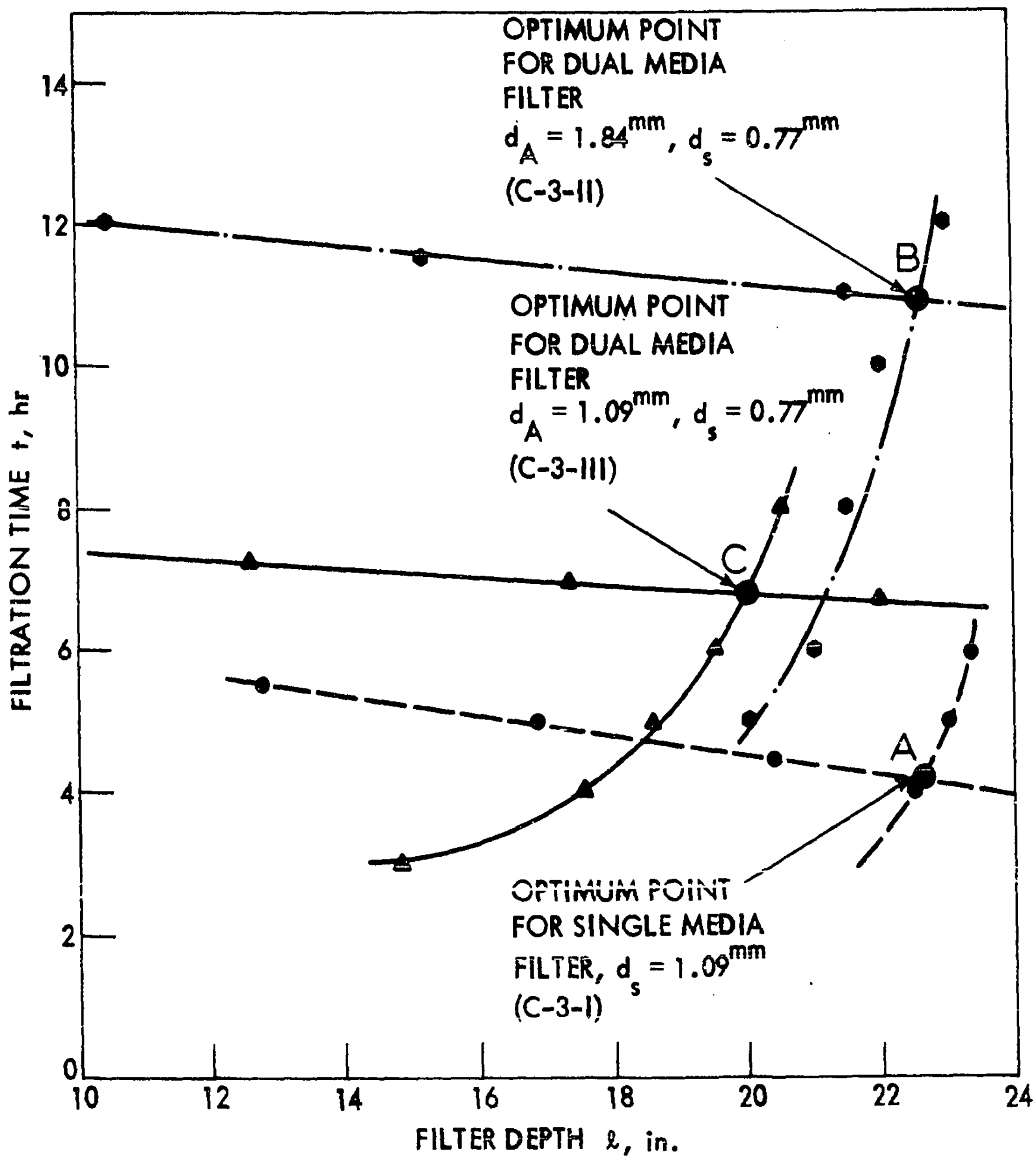

Fig. 15. Curves of values of depths and times which meet criteria C_c limit and H_c limit, for a filter run of 4 gpm/sq. ft. C-3-I, II, III

1.09 mm anthracite - 0.77 mm sand dual media filter are 22.6, 22.5 and 20.0 in., respectively. The optimum run length would be 4.2, 10.9 and 6.8 hr., respectively.

The relative operating efficiency of filters under various design conditions can be compared based on the net water production in a given period, say one day, which is defined as the result of the water produced minus the water used for backwash. A filter with shorter run lengths results in greater frequency of backwash, and greater quantities of washwater will be used during a one-day period. A filter with a longer run length results in a lower frequency of backwash, and smaller quantities of washwater will be used for backwashing in one day. It is clear that the net water production of a filter depends on the run length and the quantity of backwash water required.

Amirtharajah (1) found that the optimum backwash rate is the rate which expands the media to a 0.7 porosity during backwashing using water alone. This optimum backwash rate is the rate which expands a uni-sized sand 75-100 per cent during backwashing; for graded sand it is the rate which expands the sand 38-49 per cent. He predicted that for a graded coal system the optimum backwash rate would be the rate which would expand the coal 20-25 per cent since the original porosity in the coal bed was 0.5-0.6. Based on Amirtharajah's thoery, the optimum backwash rate for an anthracite-sand dual media filter would be the rate which would expand the media about 30-35 per cent. However, Camp, et al. (8) proposed as design criteria for backwashing dual media filters a backwash rate which expanded the media 20 per cent preceded by air scouring. Their results were based on actual field studies. Therefore, a 20 per cent expansion criteria is recommended in this study.

A summary of optimum operating conditions for single- and dual media filters are included in Table 10. The water productions for the dual media filters were 5,040 and 5,200 gallons per square foot per day; while for the single media sand filter production was only 4,140 gallons per square foot per day.

The following conclusions can be drawn based on this phase of the study:

(1) The headloss in the coarser dual media filter is only a quarter of that developed in the single media sand filter.

(2) Dual media filters have a longer optimum run length to a given headloss (10 feet of water), Table 10.

(3) A clean effluent is produced in the dual filter bed.

(4) Dual media filters require less wash water (5.3 and 6.1 per cent of filtrate in a given run against

Table 10. Optimum operating conditions for the filters in phase C, Run C-3[a]

Filter type	media depth: anthracite size mm	media depth: anthracite depth in	media depth: sand size mm	media depth: sand depth in	Optimum run length hr	Water prod. per run 1000 gal /sq.ft.	Optimum Backwash water per run: Rate gpm /sq.ft.	Optimum Backwash water per run: Quant* 1000 gal /sq.ft.	Per cent of water filtrate %	Number of runs /day**	Net water prod. per day 1000 gal /sq.ft.
Single media			1.09	22.6	4.2	1.01	40	0.20	19.8	5.10	4.14
Dual media	1.09	12.0	0.77	8.0	6.8	1.63	20	0.10	6.1	3.29	5.04
Dual media	1.84	12.0	0.77	10.5	10.9	2.62	27	0.14	5.3	2.10	5.20

* 2 minute air wash with 3-5 scfm/sq. ft. followed by 3-5 minute water wash.

** Assume 30 minute downtime per run.

[a]Run No. C-3, v = 4 gpm/sq. ft., C_o = 45.5 mg/l SS (22 JTU), T = 10.5°C, limiting filtrate quality = 8 mg/l SS, limiting headloss $\cong$ 10 ft. of water.

19.8 per cent in the single media sand filter).

(5) There is a higher net water production in the dual media filter.

Since the dual media filter has been demonstrated to be superior to the single media filter, further investigations, which will be presented in later sections, are concentrated on the anthracite-sand dual media filters for tertiary wastewater treatment.

D. Phases D & E: Anthracite-Sand Dual Media Filters

In designing an anthracite-sand dual media filter for tertiary wastewater treatment, the first question encountered will be that of what size and depth of anthracite should be used in the upper portion of the filter bed to accomplish the expected degree of suspended solids removal. A similar question applies to the sand media in the bottom portion of the filter bed. In order to determine this, a series of tests with uni-sized anthracite of various sizes on top of uni-sized sand of various sizes was conducted.

1. Selection of anthracite size and depth

a. Anthracite size   The phase C runs made with filter sets II and III provide an opportunity to examine the effect of anthracite media size on filter performance. In the phase C runs, filter sets II and III were identical (Tables 6 and 9) except that 1.84 mm anthracite was used in filter set II and 1.09 mm anthracite was used in filter set III. A total of five runs with flow rates ranging from 2 gpm/sq. ft. to 6 gpm/sq. ft. were made. Typical runs with flow rates of 2, 4, and 6 gpm/sq. ft. are summarized in Table 11.

In evaluating the results obtained with dual media filters, special care must be taken to evaluate the effects of the zone of intermixing of the two media. It should be

Table 11. Physical characteristics and operating conditions of dual media filters - varying media size of anthracite

Run no.	Filter bed: anthracite size mm	anthracite depth in.	sand size mm	sand depth in.	Media inter-mixing in.	C_o, SS mg/ℓ	C*,SS mg/ℓ	Flow rate gpm/sq.ft.
C-1-II	1.84	12	0.77	12	3	33.5	5.5	4
C-1-III	1.09	12	0.77	12	1		5	
C-2-II	1.84	12	0.77	12	2	33.8	3.5	2
C-2-III	1.09	12	0.77	12	1		3.5	
C-3-II	1.84	12	0.77	12	2	46.6	6.7	4
C-3-III	1.09	12	0.77	12	1.5		5.5	
C-4-II	1.84	12	0.77	12	2	26.0	2.2	6
C-4-III	1.09	12	0.77	12	1		2.5	
C-5-II	1.84	12	0.77	12	2	38.3	5	6
C-5-III	1.09	12	0.77	12	1		5	

* Representative filtrate quality from 4th filter cell (24 in. depth) of each filter set.

recognized that the filtering characteristics of the intermixed media zone will be significantly different from the filtering characteristics of either of the media used alone. Therefore, dual media filter designs involve selecting the size and depth combinations of both the anthracite and the sand, taking into account the effect of intermixing. If the size ratio at the interface between the sand and the coal is greater than about 3, there will be significant mixing at the interface during washing. The larger the size ratio, the greater will be the mixing of the two media.

The effects of the media intermixing at the interface of anthracite-sand after backwashing is still a subject of debate among various workers. Conley (10) reported favorable results due to intermixing and claimed that it had a very favorable influence on headloss because the finer sand grains cannot form an impervious mat when mixed with coarse anthracite. Camp (6), however, believed that the sand effected a higher degree of suspended solids removal if the top sand was not mixed with the coarser anthracite at the bottom of the anthracite coal. He suspected that Conley's excellent removal of turbidity obtained at Hanford results primarily from the use of filter-conditioning chemicals and close chemical control, and was obtained in spite of the mixing of sand and coal at the interface. It is, however, somewhat difficult to isolate the true influence of mixing alone. The degrees of intermixing experienced with the filter media size used in this study are shown in Table 11.

Fig. 16 shows the solids removal efficiency through the filter bed in three separate runs at flow rates of 2, 4 and 6 gpm/sq. ft. after 5 hours of filtration. Comparisons should be made only between the different media in a single run and not between the results of runs at different flow rates since different influent SS conditions existed in this series. In Fig. 16, it will be noted that, in filters of identical depth after the same period of filtration, the upper portion of the bed of the finer anthracite had a higher removal efficiency than that of the coarser anthracite.

In the lower portion of the 0.77 mm sand filter media (below the zone of intermixing), in all runs, there did not appear to be any increased solids removal by the additional sand and there was not a significant difference between the qualities of the effluent from the 1.84 and 1.09 mm anthracite, although finer anthracite tended to have the better quality. The same observation can be made from the three sets of data in Fig. 16 for runs at flow rates of 2, 4, and 6 gpm/sq. ft.

However, there was a significant difference in the head-loss development when different anthracite sizes were used

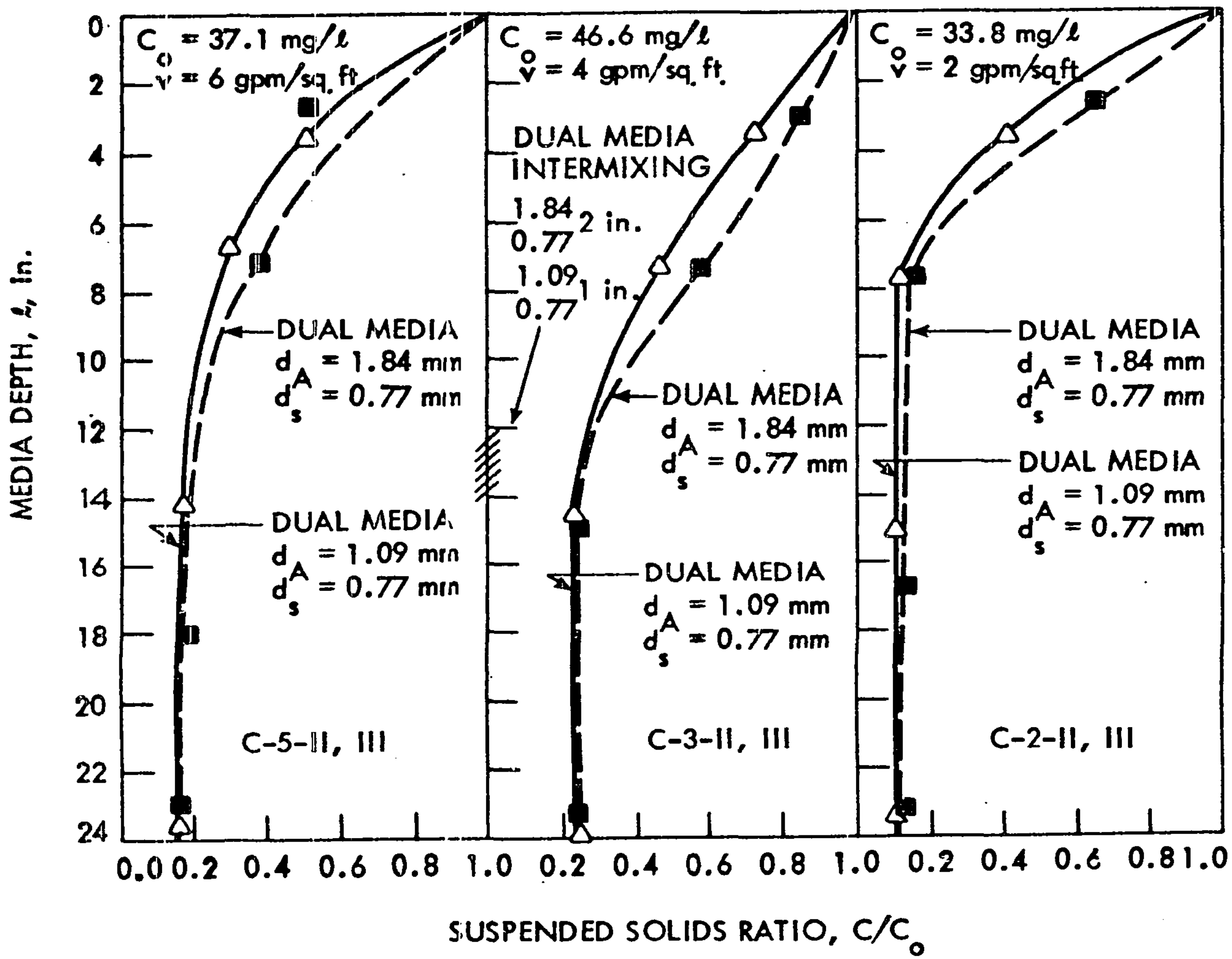

Fig. 16. Comparison of removal efficiencies of various anthracite sizes at different flow rates. t = 5 hr.

on top of the same sand size. The development of headloss with time for filters of various anthracite sizes are presented in Fig. 17. The most significant observation to be made from these data is that for each anthracite size the relationship between headloss and time is represented by an exponential curve. The removal appears not to be an adsorption removal around the media, but appears to be a bridging and/or straining, which occupies pores. Thus, the deposit is subject to compression with consequent porosity change and increasing rate of headloss with time. The rate of curvature of the headloss-time relationship is proportional to the flow rate applied, i.e. highest for v = 6 gpm/sq. ft. The intercept on the ordinate represents the clear water headloss through the filter for the stated flow rate. As indicated in Fig. 17, filters with the coarser anthracite result in lower headloss development as filtration progresses. The discrepancy increases as filtration progresses. It indicates that the 1.84 mm anthracite distributed the solids removed more uniformly throughout the filter bed.

Due to the fact that the coarser anthracite size offers significant advantages in lower headloss build-up and no significant disadvantages as far as filtrate quality is concerned, a uni-sized 1.84 mm anthracite was used for further investigation.

b. Anthracite depth Another conclusion suggested

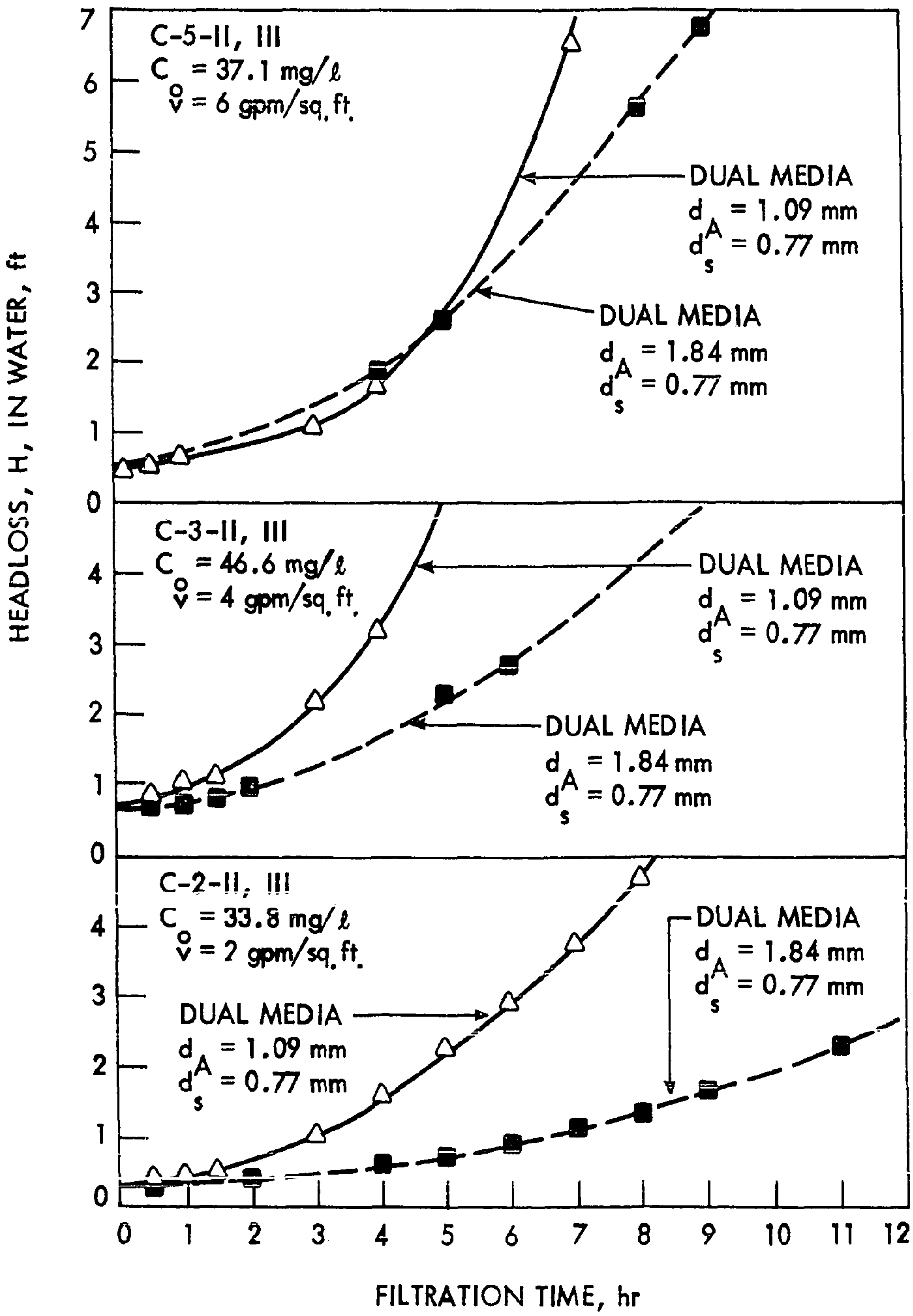

Fig. 17. Comparison of headlosses of various anthracite sizes at different flow rates through the 24 in. full depth filter

by the data in Fig. 16 was that there was no significant further removal beyond a filter bed of 12 inches (or below the media intermixing zone) regardless of the flow rate.

The same conclusion concerning the adequate anthracite depth can be drawn from headloss information. Curves in Fig. 18 indicate that the headloss build-up occurs mostly in the upper few inches of the filter bed and its rate of build-up (H/l) increases as the size of anthracite becomes smaller. Also, no significant headloss is generated beyond the 12 inch anthracite depth. This is in accordance with the fact shown in Fig. 16, that no further significant removal occurs below a depth of 12 inches.

The conclusion drawn from the data shown in Figs. 16 and 18 was based on the results of various degrees of intermixing in the filter bed after backwashing. It is important to investigate the removal efficiency through the filter bed which contains anthracite alone and thus no intermixing. By doing this, the necessary depth of anthracite above the intermixing zone can be determined.

With this purpose, phase E of this study was conducted to investigate the effect of media size on filtration of wastewater through single media anthracite filters. Filter characteristics of this phase were as follows:

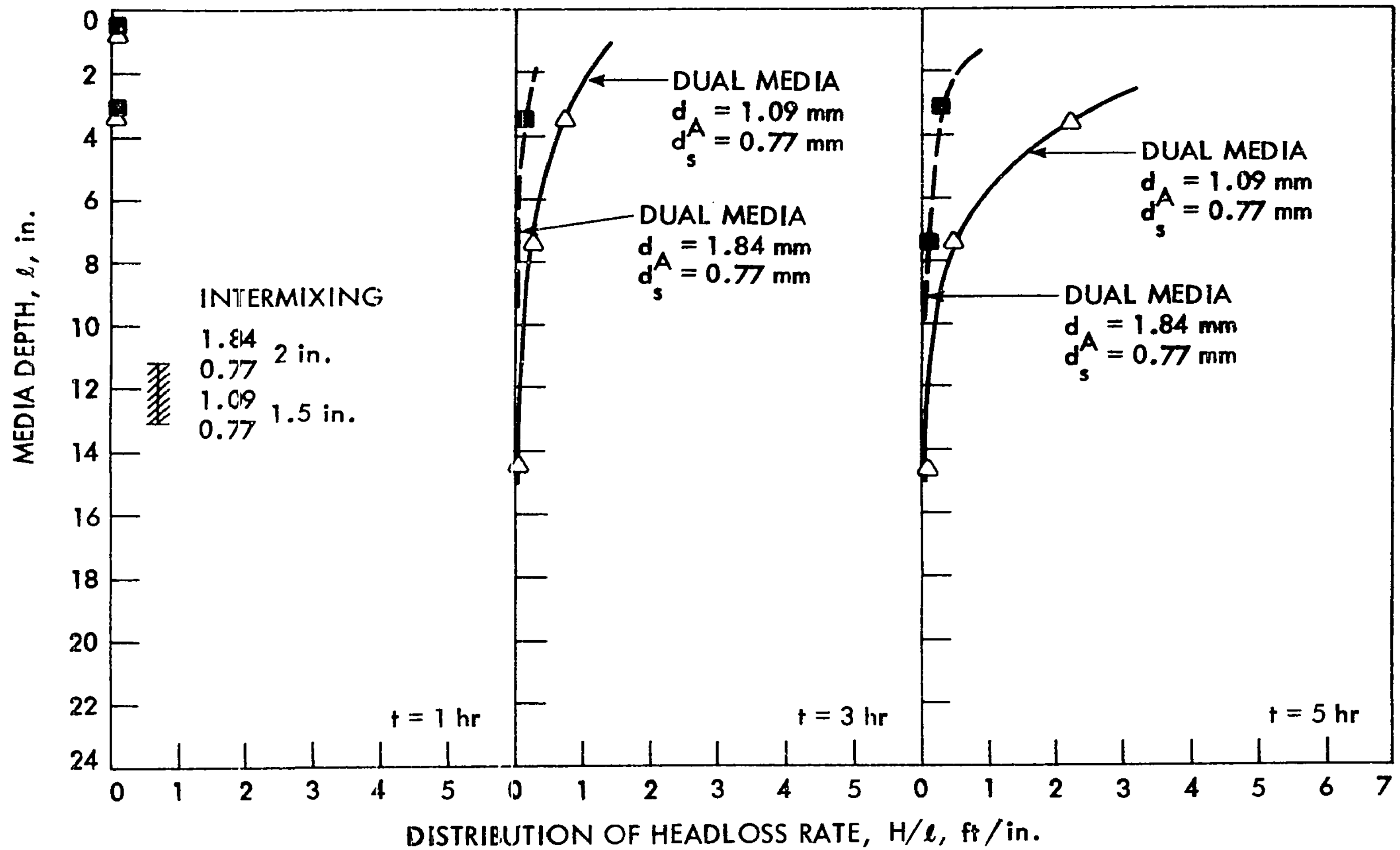

Fig. 18. Comparison of distribution of headloss rate for various anthracite sizes at a flow rate of 4 gpm/sq. ft., C-3-II, III

Filter set I - 24 in. of uni-sized 1.85 mm anthracite

Filter set II - 24 in. of uni-sized 1.30 mm anthracite

Filter set III - 24 in. of uni-sized 1.09 mm anthracite

A total of three runs with flow rates of 2, 4 and 6 gpm/sq. ft. was made, with results in Fig. 19.

As shown in Fig. 19, no further solids removal for all sizes below the anthracite depth of 8 in., 10 in., and 10-12 in., respectively was observed at flow rates of 2, 4 and 6 gpm/sq. ft. respectively. It again indicates that the extent of solids penetration depends on the flow rate applied. It can be concluded that maximum anthracite depth required in dual media filter design would be around 12-15 in. on top of the sand.

2. Selection of sand size and depth beneath 1.84 mm anthracite

a. Sand size Dual media filters consisting of 12-15 inches 1.84 mm anthracite on top of sand have been demonstrated to be a reasonable design which can be operated successfully under various flow rates and the Ames final wastewater characteristics. The task of selecting the proper sand size and depth for the bottom portion is no less important.

In view of the fact that the filter design used in this study provided for a maximum filter media depth of about 24 inches, all studies of the effect of sand media size were conducted using a 12 inch depth of 1.84 mm anthracite as the top media. Thus, the three filter sets were provided

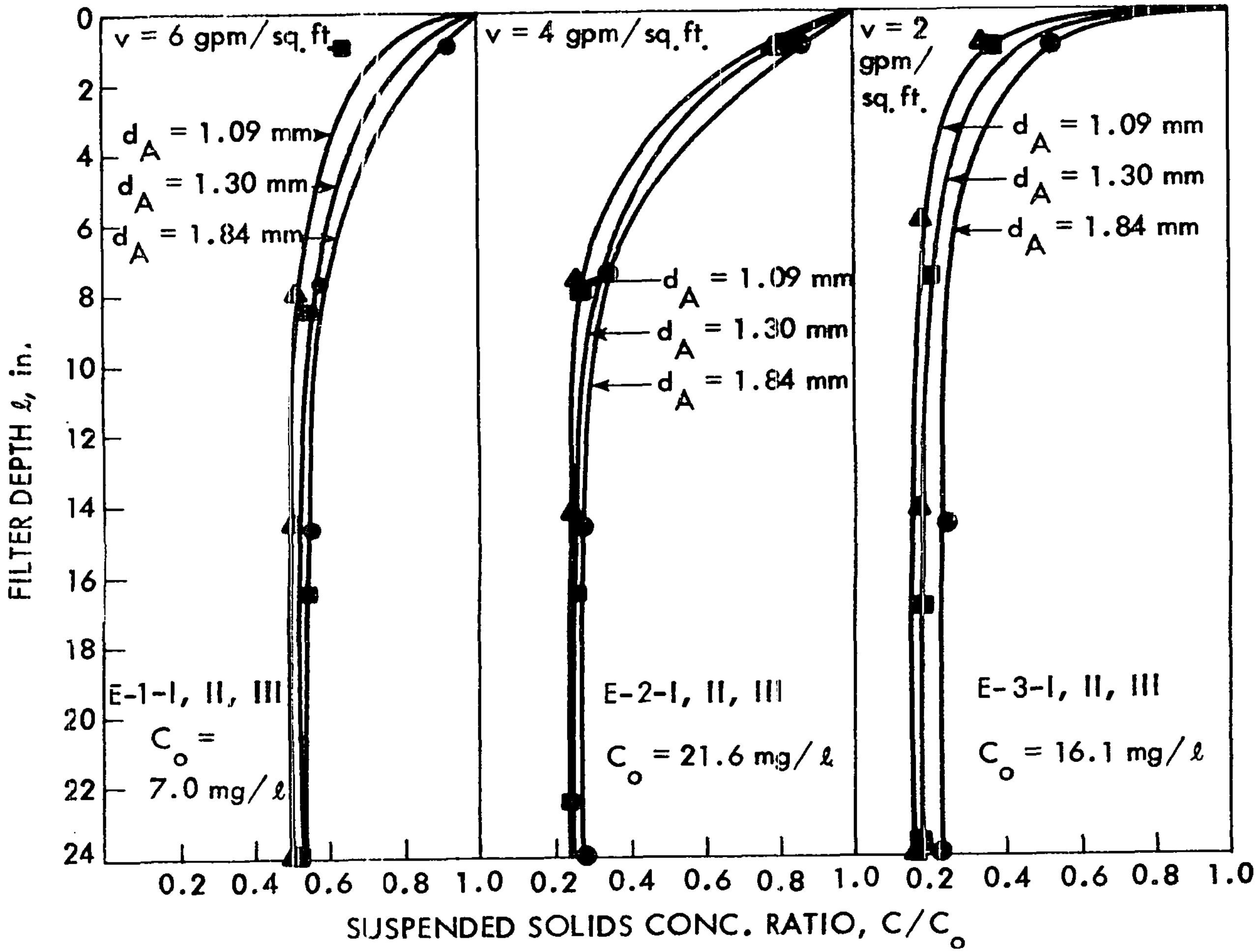

Fig. 19. Effect of single media anthracite size on solid removal efficiency at t = 11 hr.

with a 12 inch depth of 0.77, 0.65, and 0.55 mm sand as shown in Table 12. The phase D studies were conducted to evaluate the effect of the sand media size on filter performance.

Three filter sets were operated in parallel under the same flow rate and applied wastewater conditions. In this arrangement, sand size was the only variable. Therefore, any difference in performance (filtrate quality and headloss development) is assumed to be contributed by this variable. A total of five runs with flow rate ranging from 2 gpm/sq. ft. to 6 gpm/sq. ft. were made (Table 12).

The solids removal efficiencies through the filter beds in three runs at flow rates of 2, 4 and 6 gpm/sq. ft. are shown in Fig. 20. It is interesting to note that the removal efficiency obtained from all three filter sets is almost identical at a depth of 24 inches at the 2 gpm/sq. ft. flow rate. There appear to be only slight variations in quality between the filters at higher flow rates (4 and 6 gpm/sq. ft.). The removal efficiencies within the top 12 inches of the filter are identical at all flow rates. This is expected as the anthracite in the top portion above the media intermix zone is the same size in all filters. The existence of a media intermix zone did not appear to dramatically affect the solids removal relationship observed.

The headloss development versus time in the same runs

Table 12. Physical characteristics and operating conditions of dual media filters - varying media size of sand (phase D runs)

Run no.	C_o, SS mg/ℓ	Rate gpm/ sq.ft.	Filtrate C*, SS mg/ℓ media sizes, mm			Headloss ft. water	Run length* to given headloss, hr. media sizes, mm		
			1.84[a] 0.77	1.84[b] 0.65	1.84[c] 0.55		1.84[a] 0.77	1.84[b] 0.65	1.84[c] 0.55
1 **									
2 **									
3	40.5	6	5.4	5.2	5.0	10	9	5.5	7
4	12.1	4	2.8	2.8	2.8	10	28	25	25
5	20.0	2	2.8	2.8	2.6	10	60	46.5	52

* Representative filtrate quality from 4th filter cell (24 in. depth) of each filter set.

** Mixer out of order.

[a] 1 in. intermixing.

[b] 3~4 in. intermixing.

[c] 4~6 in. intermixing.

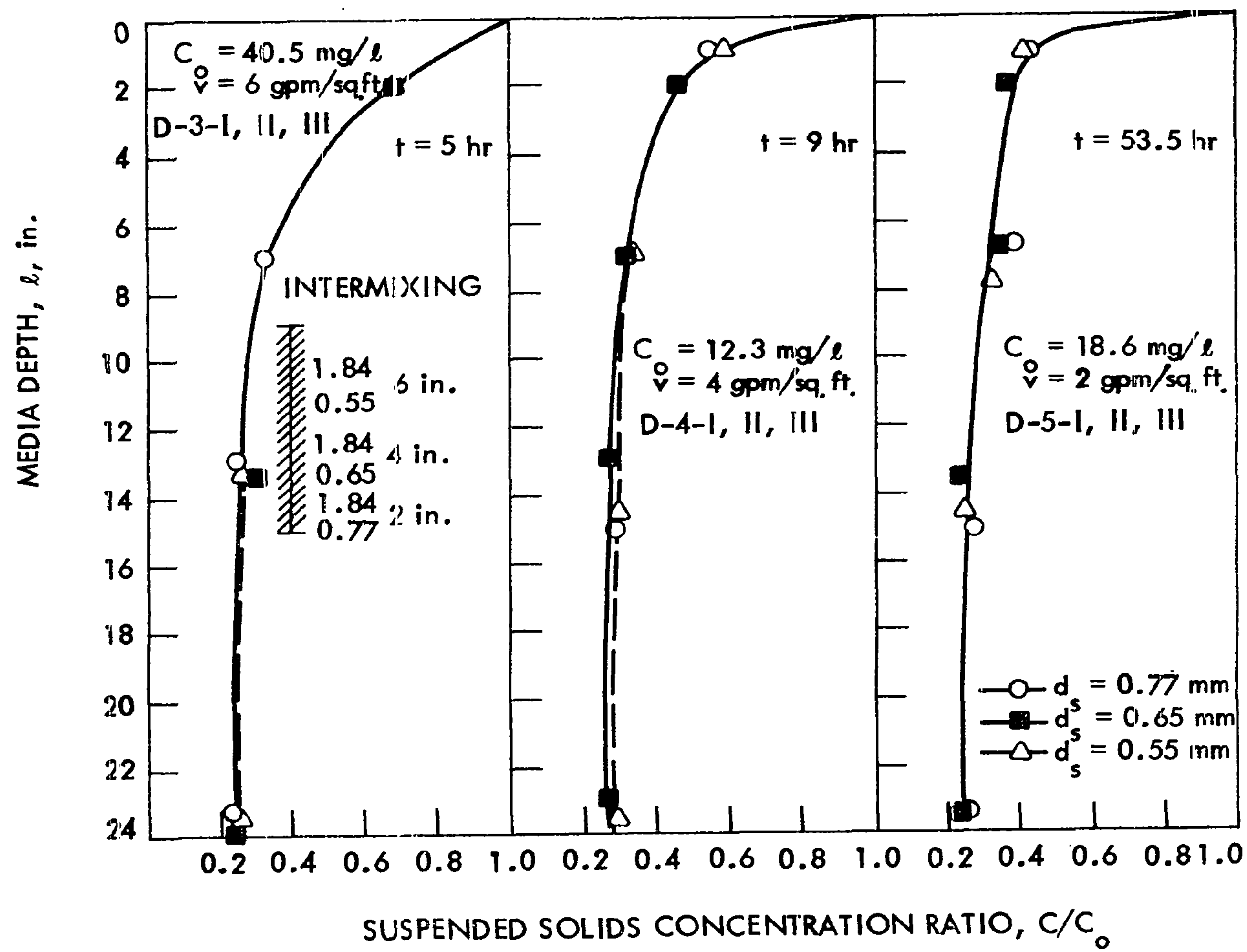

Fig. 20. Comparison of removal efficiencies of various sand sizes at different flow rates

is shown in Fig. 21. As expected, the filter with the coarsest sand (0.77 mm) beneath the anthracite produces a slightly lower headloss development as filtration progresses. However, the filter with the finest sand (0.55 mm) used in this investigation shows a lower headloss development versus time than that of the filter with the 0.65 mm sand. These results suggest the conclusion that a filter with coarse anthracite (1.84 mm) on top of fine sand (0.55 mm) can produce a slightly better effluent quality without having a greater headloss development than those filters using the same anthracite layer above a coarser sand size. This size combination results in a longer run length. This might be due to substantial intermixing at the interface after backwashing of the dual media. A 6-inch intermixing was generally observed for this kind of size combination.

b. Sand depth The selection of sand depth in a dual media filter remains more debatable. In general practice, sand depth used has ranged from 6 to 30 in. with 8 to 12 in. most common (6, 10). So far, no rational method is available to determine the sand depth. Experimental data shown in Fig. 20, which show the suspended solids removal efficiencies through the filter bed when the sand size in the bottom portion was the only variable, indicates no further solids removal occurred beyond the media depth of 12 in. This indicates a sand depth of 12 to 15 in. beneath 12 to 15 in. of anthracite

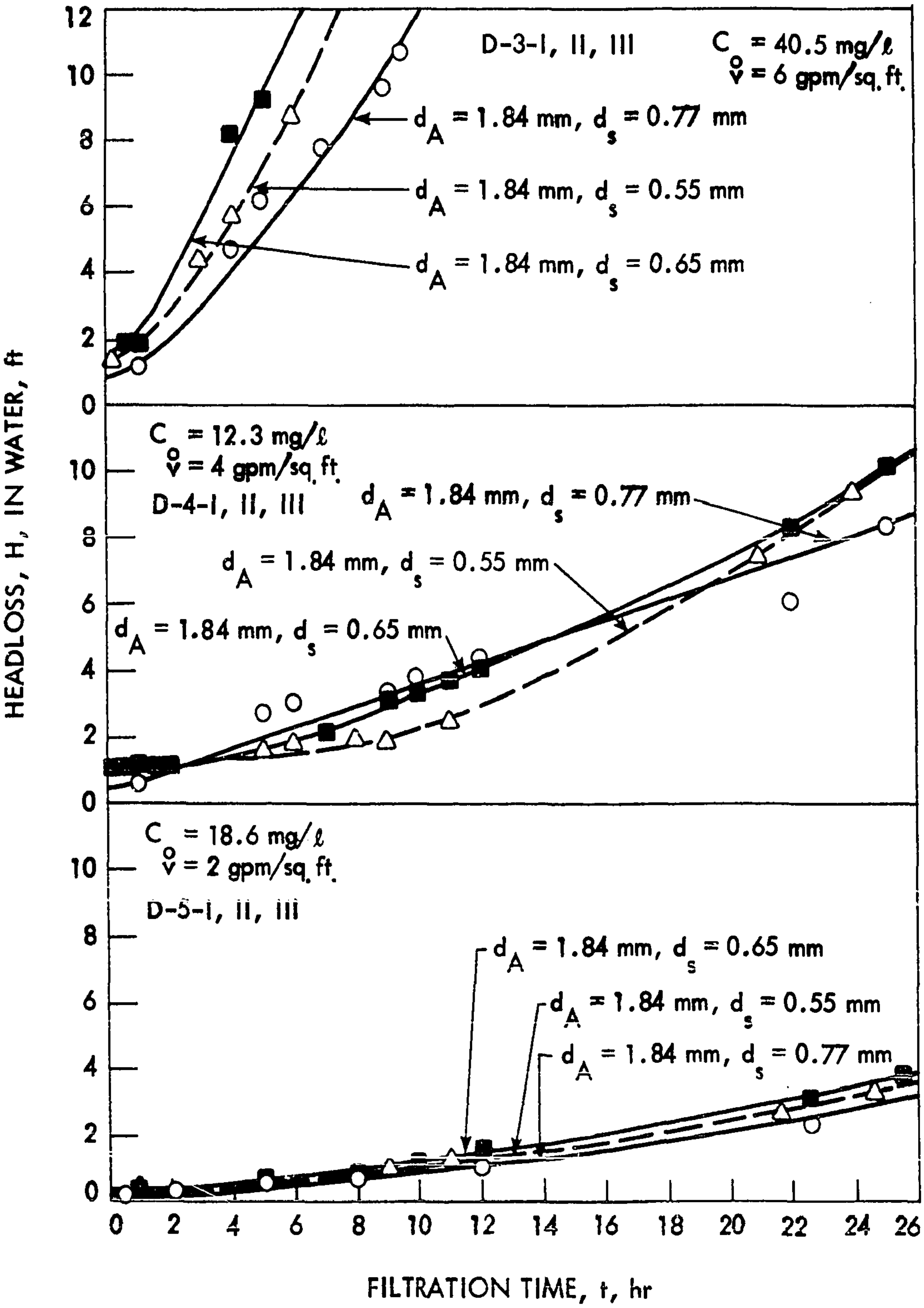

Fig. 21. Comparison of headlosses in dual media filters using various sand sizes at various flow rates

is sufficient.

E. Phase F: Flow Rate Effects on Dual Media Filters

On the basis of the tests completed in phases A, B, C, D, and E, it has been determined that for the filtration of the Ames final effluent wastewater a dual media filter should have the following characteristics:

Anthracite size - uni-sized 1.84 mm (10/12)

Anthracite depth - 12 to 15 in.

Sand size - 0.55 mm (30/35)

Sand depth - 12 to 15 in.

In view of this, the phase F studies were conducted using filter sets which were identical in a series of runs to evaluate the effect of flow rate on filter performance. A summary of runs is presented in Table 13 and typical data resulting from these experiments are presented in the following sections.

1. Effect on SS removal efficiency

The curves shown in Fig. 22 represent the normalized change in solids concentration (the ratio of effluent to influent suspended solids) achieved with depth for three different flow rates in the 1.84 mm anthracite - 0.55 mm sand dual media filter.

Table 13. Summary of results of phase F runs

Run no.	C_o, SS mg/ℓ	Media size mm	Filtrate C*, SS mg/ℓ Flow rate, gpm/sq.ft. 2	4	6	Headloss ft. water	Run length* to given headloss, hr. Flow rate, gpm/sq.ft. 2	4	6
1	12.5	1.84 0.55	0.5	0.9	1.1	10	55	32	18
2	12.9	1.84 0.55	0.6	0.9	1.8	10	53	27	19

* Representative filtrate quality from 4th filter cell (24 in. depth) of each filter set.

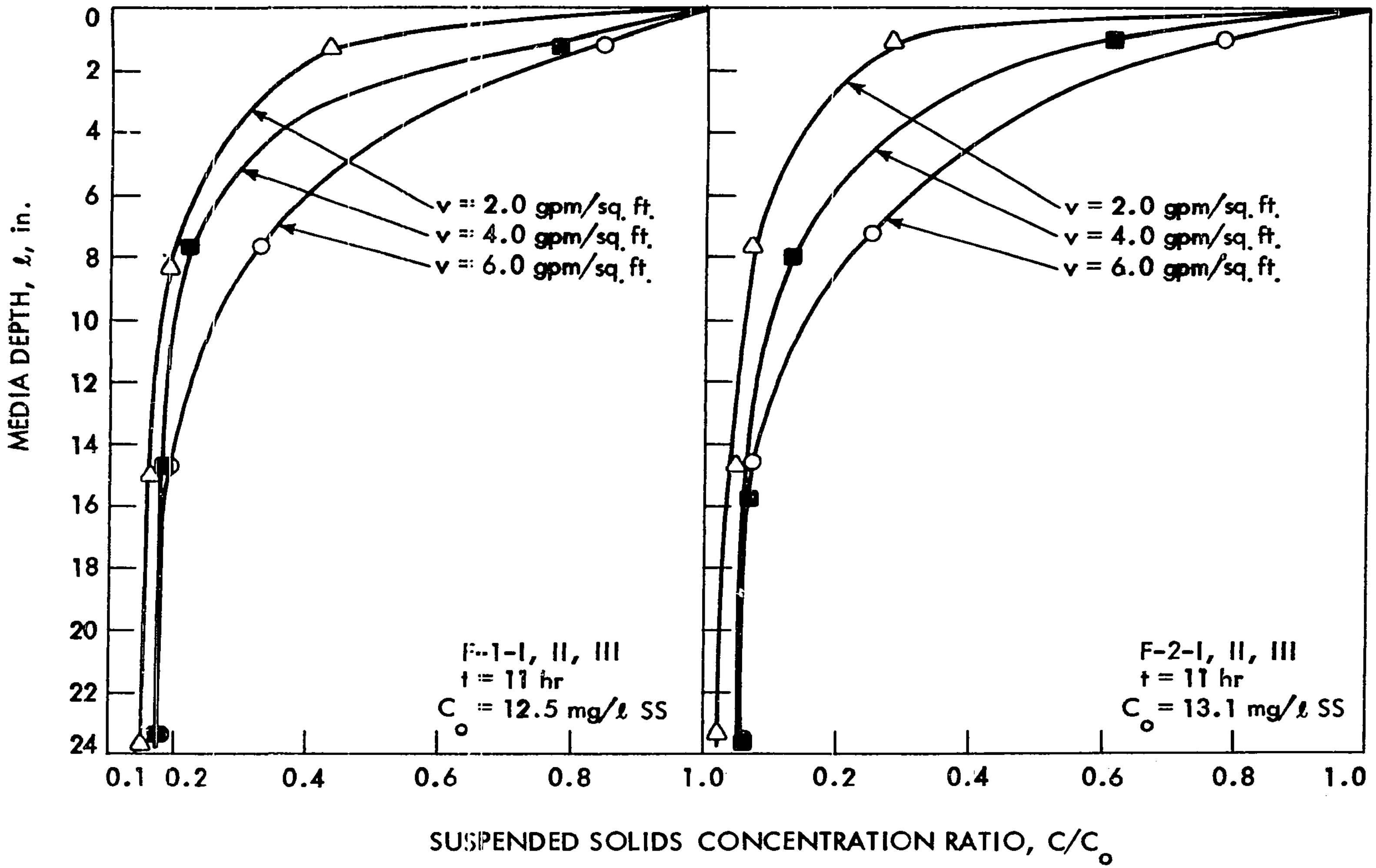

Fig. 22. Comparison of removal efficiencies of various flow rates (12 in. - 1.84 mm anthracite on top of 12 in. - 0.55 mm sand)

Fig. 22 indicates that the suspended solids removal efficiency is lower in the top portion of the bed at higher rates of filtration. The higher flow rate causes deeper penetration of the suspended solids. Apparently, the larger shear forces due to the high flow-through velocity at the high rate of filtration reduce the removal in the upper portion of the bed. As evidenced in these curves, flow rates of 4 and 6 gpm/sq. ft. have the same removal efficiency where the media depths are both 24 in.

It was found that effluent quality from the media depth of 24 in. was not significantly affected by a flow rate up to 6 gpm/sq. ft., as shown in Fig. 23. This can be further demonstrated by the curve in Fig. 24, which shows the relationship of effluent quality ratio to the flow rates at a filtration time of 22 hr. The line in Fig. 24 shows a nearly horizontal relationship which indicates little effect on effluent quality due to the increase of flow rate. This conforms to the results of Tebbutt (86) and Yao, et al. (96).

The following additional observations can be drawn from Fig. 23:

(1) Effluent quality was more sensitive to the influent quality during the early stages of the filter run, but was not affected by influent quality during the latter part of a run.

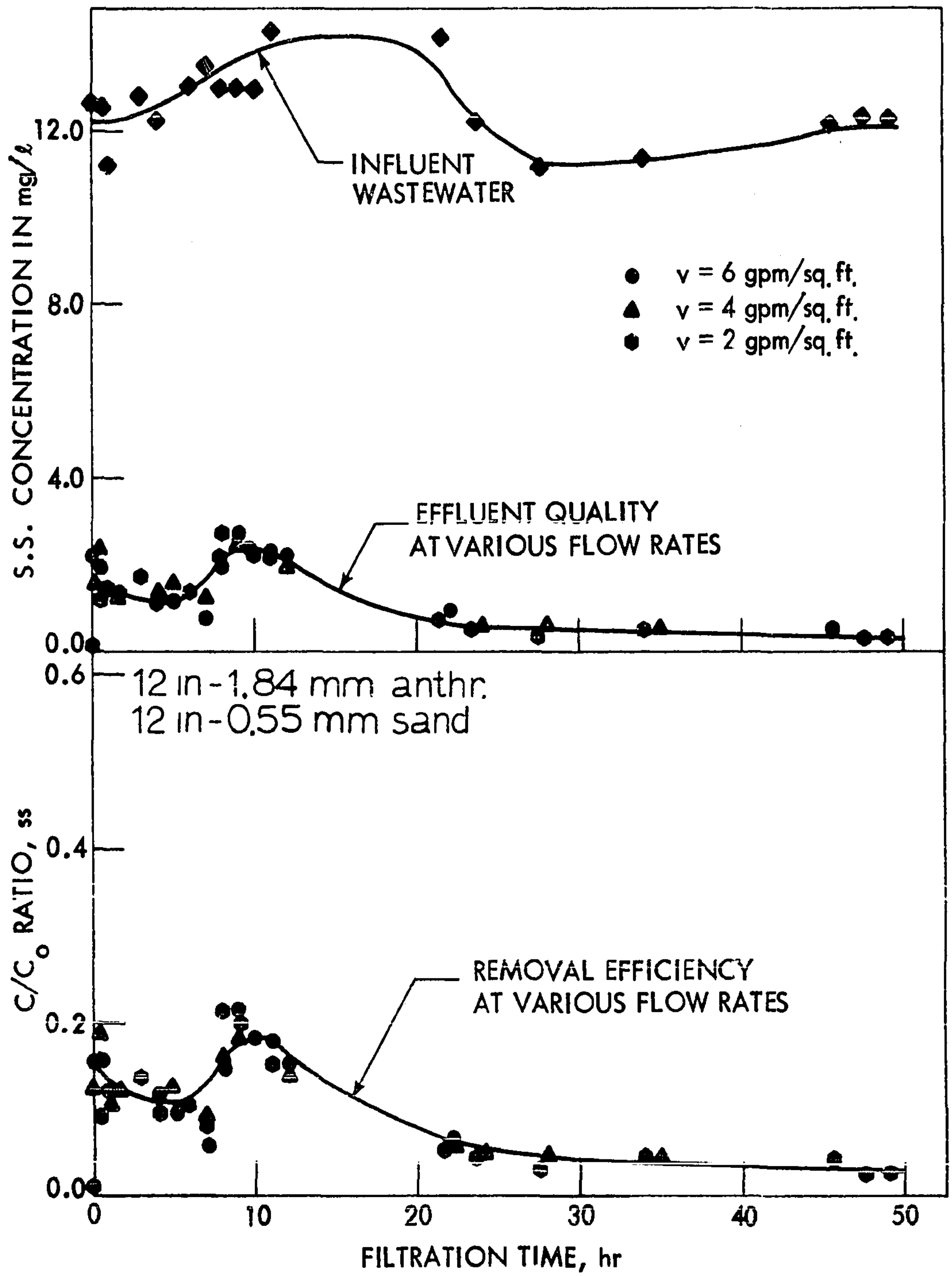

Fig. 23. Comparison of effluent qualities and removal efficiencies at various flow rates, F-1

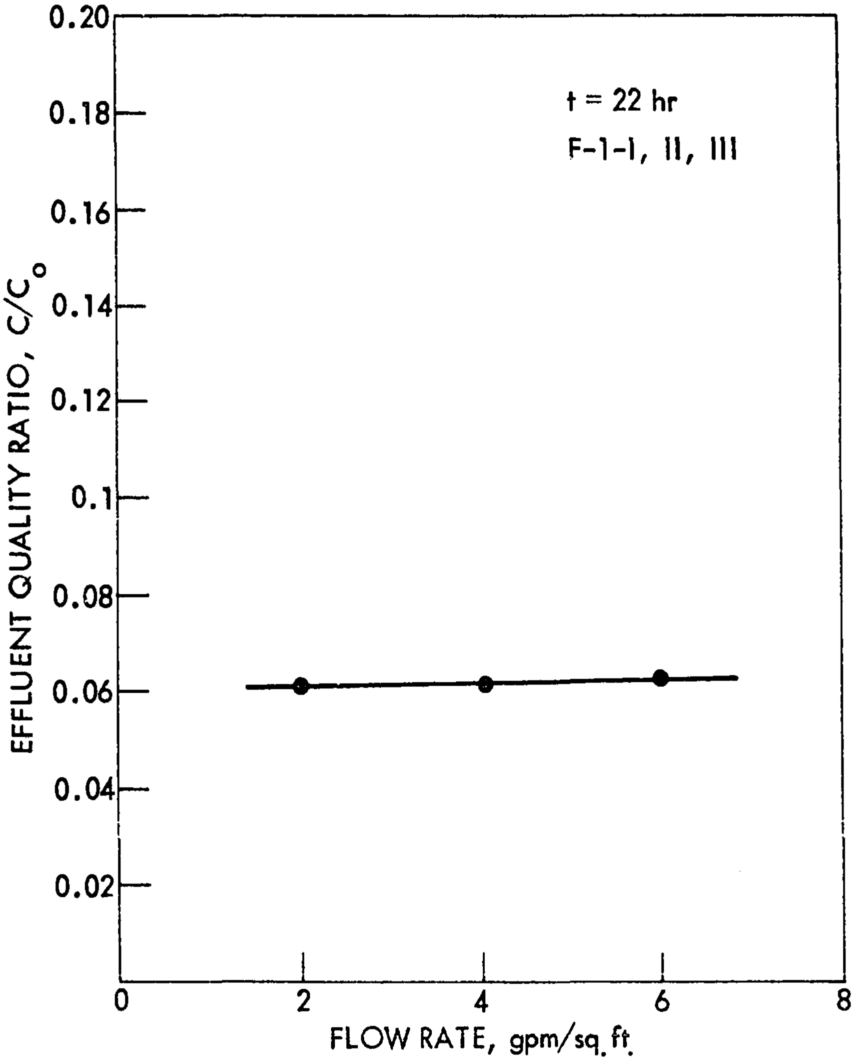

Fig. 24. Effluent quality ratio vs. flow rate at filtration time of 22 hr. F-1-I, II, III

(2) Over 90 per cent SS removal was obtained regardless of the flow rates.

(3) No sign of breakthrough occurred even 55 hours after the beginning of a run.

The finding that a higher flow rate need not deteriorate the effluent quality is encouraging, since it means more economical filter operating conditions can be obtained. If the flow rate is nominally doubled, the area of the filters can be approximately halved, which will result in some saving.

2. Effect on run length

Fig. 25 shows the progress of headloss development at various filtration times. As shown in the figure, higher headloss results from higher flow rate under the same filtration time. This might be because at a higher flow rate more wastewater was filtered during the same amount of time than would have been filtered at a lower flow rate, thus resulting in a higher solids accumulation within the filter pores. If the maximum allowable headloss is fixed at 10 ft. of water, the run lengths for the filters at flow rates of 6, 4 and 2 gpm/sq. ft. are 18, 32 and 55 hrs., respectively. Since no sign of suspended solids breakthrough occurred even as long as 55 hours after the beginning of the run, it is obvious that the headloss limit available controls the run length, which is very much dependent on the flow rate.

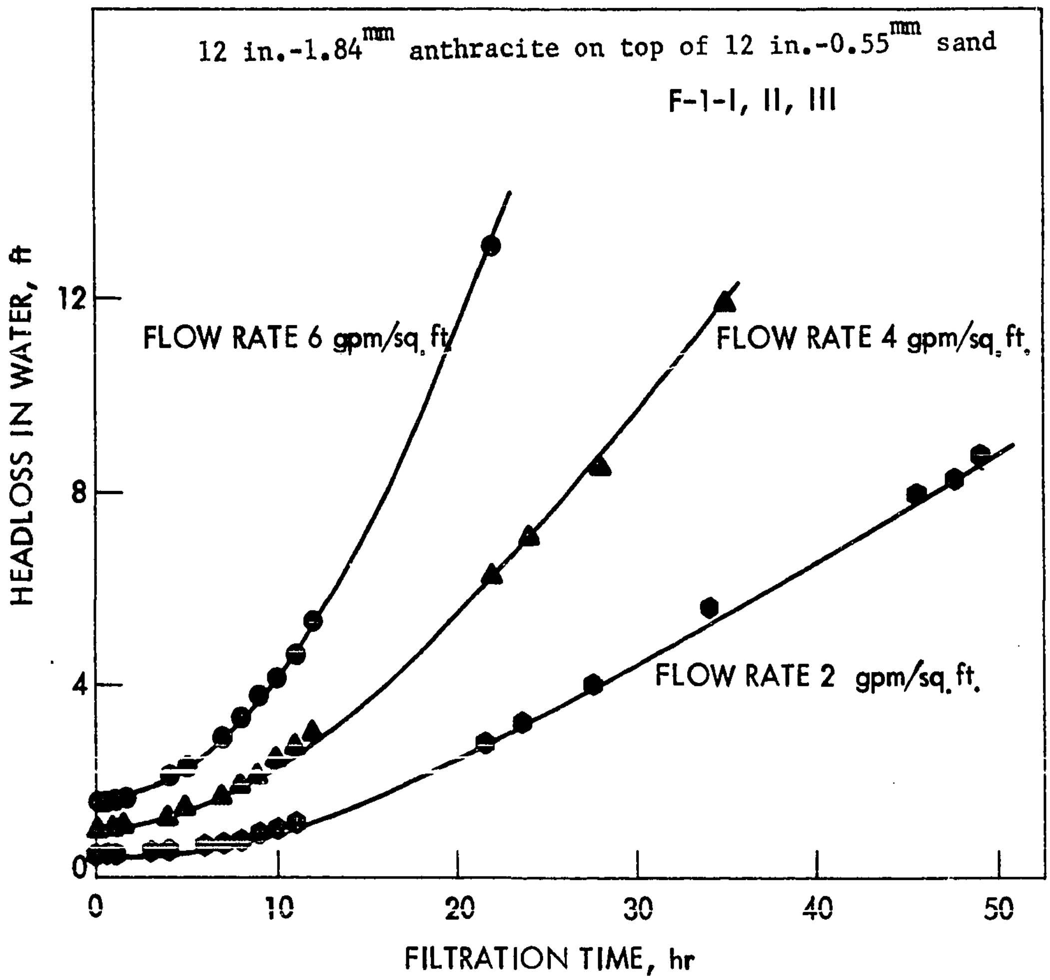

Fig. 25. Comparison of headloss vs. time at various flow rates in Phase F run

F. Chemical Removal in Filtration

Plain filtration of wastewater removes SS, BOD and other chemical pollutants. It is believed that the BOD removed by filtration is primarily the portion associated with the suspended solids, that is, the portion of undissolved BOD. The amount represented by the undissolved BOD varies greatly with the type of wastewater. Furthermore, the portion of undissolved BOD from the final effluent of a trickling filter plant is expected to be different from that of an activated sludge plant, due to the difference in degree of cell synthesis.

Table 14 shows a summary of the removal of chemical pollutants in a typical Phase C filter run conducted at a filtration rate of 6 gpm/sq. ft. The following observations were drawn from Table 14.

(1) Filtration results in a high suspended solids removal: 87 per cent for the 1.09 mm sand filter and 88.5 per cent for both anthracite-sand dual media filters.

(2) BOD per cent removals due to filtration ranged from 43 to 46 per cent. A slightly better result was obtained from the dual media filters.

(3) COD per cent removals due to filtration ranged from 36.4 to 37.6 per cent with slightly better results from the dual media filter.

Table 14. Removal of chemical pollutants during a typical filter run (C-5), flow rate: 6 gpm/sq. ft.

Chemical constituent	Final effluent concen-tration mg*/ℓ	Filter Set I 24 in. - uniform 1.09 mm sand		Filter Set II 12 in.-1.84 mm anthr. 12 in.-0.75 mm sand		Filter Set III 12 in.-1.09 mm anthr. 12 in.-0.75 mm sand	
		Filtrate quality mg/ℓ	Removal efficiency %	Filtrate quality mg/ℓ	Removal efficiency %	Filtrate quality mg/ℓ	Removal efficiency %
Turbidity	20 JTU (0-100 scale)	7.5 JTU (0-10 scale)		6.7 JTU (0-10 scale)		6.4 JTU (0-10 scale)	
SS	40	5.2	87.0	4.9	87.8	4.6	88.5
BOD	28.0	16.0	42.9	15.8	43.6	15.2	45.7
COD	77.0	49.0	36.4	48.0	37.6	48.0	37.6
TOC	19.0	13.0	31.6	12.0	36.8	12.0	36.8
Ortho-PO_4	18.9	18.9	0	19.1	0	18.6	0
total-PO_4	20.5	19.6	4.4	19.9	2.9	18.9	7.8
Org-N	12.2	9.4	23.0	9.3	23.8	11.2	8.2
NH_4-N	10.1	9.2	8.9	9.5	5.9	9.4	6.9
NO_2-N	0.88	0.83	--	1.15	--	0.96	--
NO_3-N	4.18	4.45	--	4.04	--	3.43	--

*Except turbidity, which is in JTU.

(4) TOC per cent removals due to filtration ranged from 31.6 to 36.8 per cent with slightly better results from the dual media filter.

(5) Removal of orthophosphate and total phosphate due to filtration was negligible.

(6) Removal of organic nitrogen due to filtration ranged from 8.2 to 23.8 per cent.

(7) No significant removal of other forms of nitrogen due to filtration was observed.

VI. PRACTICAL ASPECTS OF WASTEWATER FILTRATION

A. Dilemma of Optimization

During a filter run using trickling filter final effluent filtrate quality and headloss development vary. Usually, the filtrate quality falls during the first hour or two, remains steady for a considerable period, and then starts to deteriorate, giving a poorer filtrate with time (Figs. 6, 7, 8, 10, and 23). Meanwhile, the headloss increases continuously, either linearly (Figs. 6 and 7) or exponentially (Figs. 6, 7, 8, 12, and 25). Both have limiting values set by quality requirements for the filtrate and by hydraulic conditions for the headloss.

Usually, a number of different filter designs may be available to meet a particular filtration requirement. Either fine media shallow filter operated at low rate or a coarser media deep filter operated at a high rate may provide the desired filtrate quality at a reasonable run length to a given headloss limit. However, it does not follow that any of the many potential filter design combinations (headloss limit, filtration rate, and media size and depth) provides a design and operation that will produce filtered water at least cost. The concept of optimum design has received increasing attention, due particularly to the work of Mintz (59, 60), Ives (35-37) and Huang and Baumann

(27). Mintz proposed the concepts of "time of protective effect (t_1)" and "time to limiting headloss (t_2)." The filtrate quality changes with time during a filter run, and the "time of protective effect" is the length of run to the moment when the filtrate quality reaches an unacceptable value. The headloss rises during the run, and the "time to limiting headloss" is the length of run to the predetermined headloss limit.

If t_1 is greater than t_2, the filter has a reserve solids removal capacity which is unutilized. It indicates that the filter media is either too deep or too fine. Conversely, if t_2 is greater than t_1 (which leads to solids breakthrough), the depth of the filter bed is too shallow or the media is too coarse. Mintz (59) included in his report a number of charts showing how optimization of filter design and operation could proceed.

The technique of optimization has been advanced further by Ives (35, 37) and Ives and Gur[1]. Ives and Gur proposed the concept of operational optimum and economic optimum. The operational optimum occurs in a filter when the clarification capacity of the filter is exhausted simultaneously with its hydraulic capacity, $t_1 = t_2$. Operational optimums

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[1]Ives, K. J. and Gur, A. University College, London, England. Research on optimization of filtration. Private communication. 1971.

depend only on the media size and depth, flow rate and hydraulic conditions limiting the headloss. The economic optimum applies to an operational optimum filter which produces filtered water at least cost. Economic optimum filters require consideration of the costs of the filter structure, hydraulic appurtenances, energy and maintenance requirements, and depend also on the size of the filter plant. They presented a graphical method together with a theoretical mathematical equation for the operational and economic optimization of filtration.

Huang and Baumann (27) have also contributed to the optimization of the filtration process. Using an empirical model developed by Hsiung and Cleasby (26), they found that there exists a set of optimum filter design variables (sand depth, flow rate, terminal headloss and run length) at each specific sand size. When the sand size changes, the set of optimum design variables changes accordingly.

It is generally agreed that two elements are essential in optimization of a filtration system. They are:

(1) A method (mathematical model) to predict the performance of the filters to determine combinations of the filter design variables which will provide the desired filtrate quality under operational optimum conditions.

(2) A filtration system whose first cost, operating

cost, and maintenance cost can be predicted with reasonable accuracy.

A computer program based on those two elements can be developed to provide a list of operational optimum filter designs and to predict the one design which will provide the treated water at least cost.

To date, no mathematical model is available which can be used successfully to predict operational optimum conditions in tertiary wastewater filtration.

All optimization models used to date have been based on the removal of iron from suspension by filtration (27, 35, 37, and 59). Iron suspensions were used to provide the empirical coefficients in the models because the concentrations of the suspensions and the characteristics of the iron are rather constant such that filter performance can be predicted with a reasonable degree of accuracy. Since similar conditions are encountered in field filtration situations, such models are of immediate practical importance. However, in the case of wastewater filtration, the wastewaters coming to the tertiary filter plant vary over a wide range in their suspended solids concentration, physico-chemical properties and the particle size distribution of the suspended solids. As a result, it is difficult to conduct pilot-plant tests with the "significant wastewater" whose solids the plant is to be optimized to remove. Once a plant is optimized, any change in the amount or character of the solids delivered to it will completely upset the filter operating characteristics and the economics of the filtration.

The inability to date to develop a mathematical model to predict tertiary treatment filter performance is due to the major variations in the quality of the wastewater coming into the filter plant. These are due primarily to the variations in treatment efficiency of the preceding units (primary, biological treatment and final clarifier). The treatment efficiency of the primary and biological treatment unit varies seasonally due to the change in temperature and even hourly due to the fluctuation of the organic load and hydraulic loading. As a result, the characteristics of the wastewater coming into a tertiary filter plant can be expected to vary seasonally and hourly.

The quality of the wastewater coming into a tertiary filter plant varies from time to time in three major ways:

(1) in the concentration of the suspended solids,

(2) in the physico-chemical properties of the suspended solids, and

(3) in the particle size distribution of the suspended solids.

This variation in influent solids characteristics affects the filter removal mechanisms in two ways:

(1) variation in the proportion of the total solids which are removed through surface removal versus depth removal mechanisms.

(2) variation in the proportion of the total solids whose removal is more affected by transport phenomena versus those whose removal is more affected by attachment phenomena.

To date, the characteristics of the final effluent have not been investigated fully. Suspended solids and turbidity, the only measurements commonly made of effluent solids concentrations give only limited information about the characteristics of the solids in the wastewater makes prediction of filter performance difficult, if not impossible.

1. Effect of variation of influent SS concentration on headloss development

The pilot filter plant, which filtered the final effluent from the Ames water pollution control plant, was not able to filter wastewaters with different degrees of influent SS concentrations simultaneously to the various filter sets during a single filter run. Therefore, this study was unable to treat influent SS concentrations, C_o, as the only variable under investigation during a single run. However, some observations can be made concerning runs with different influent SS concentrations with respect to the effect on the headloss development due to the difference in SS concentration.

According to Ives (33), the headloss through a filter bed is related to the volume of the deposited flocs. Therefore, for a specific media and flow rate, the total headloss depends only on the volume of solids retained by the filter. This leads to the conclusion that, for a given filter at a constant flow rate, the time needed to reach a fixed headloss depends only on the volume of the filter pores filled with the suspended solids removed from the raw water. In his recent work, Ives (35) proposes integration of the headloss with media depth to account for the distribution of the SS along the filter bed.

The determination of the volume of floc removed in filtration of secondary final effluents would be of little interest and difficult to reproduce. Therefore, Tchobanoglous and Eliassen (85) proposed an empirical equation relating the development of headloss to the amount of solids accumulated within the filter pores. They assumed that headloss (Equation 29) was dependent only on the amount of the solids accumulated within the filter pores. These runs were made with a constant flow rate and a fairly uniform suspended solids level and characteristic, so no general conclusions could be drawn from them about the effect of flow rates or suspended solids level or their characteristics.

However, data from this pilot plant study shows that headloss is affected by other variables, such as the influent SS concentration and/or particle size distribution in the wastewater and flow rate as well as the amount of solids accumulated within the filter pores. Fig. 26 shows the headloss development observed due to the accumulation of solids within the filter pores at several influent SS concentrations. Three cases are shown at each of three flow rates, 2, 4, and 6 gpm/sq. ft. Two obvious observations are drawn from Fig. 26.

(1) There is a variation in headloss development between runs conducted at the same flow rate even though the solids accumulation within the filter pores is the same. This variation shows a random nature, but does not show a definite functional relationship of C_o.

(2) The variation is least at a flow rate of 2 gpm/sq. ft. and most at a flow rate of 6 gpm/sq. ft.

The explanation for this variation might be that there is a variation in the SS characteristics (physico-chemical properties and/or particle size distribution) even with the same C_o levels. This variation in the SS characteristics might affect the degree of solids distribution through the filter bed, which would cause the difference in headloss development observed.

Two sets of runs whose data are included in Fig. 26

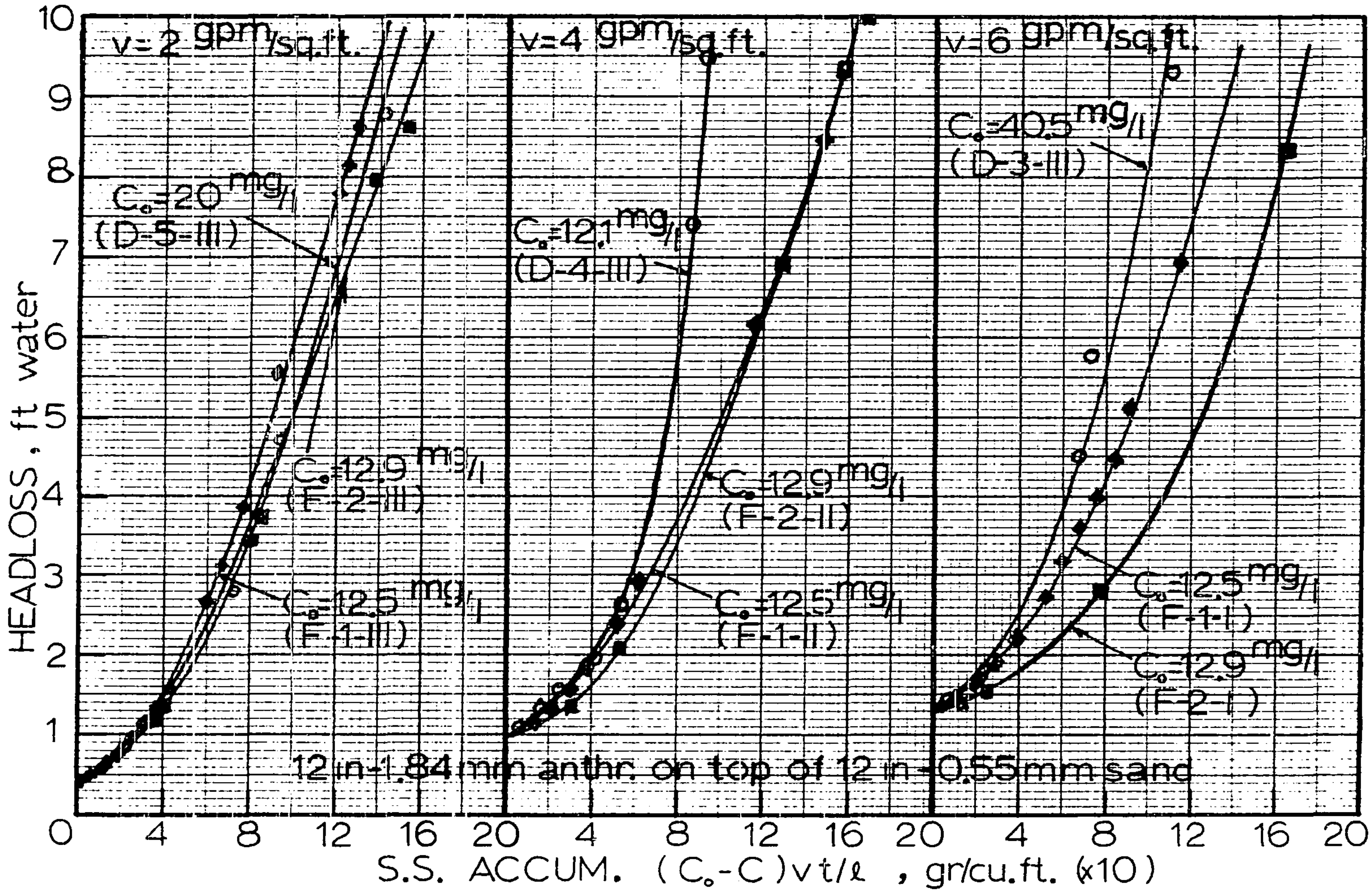


Fig. 26. Headloss vs. SS accumulation at various levels of SS concentration in wastewater

were made so that data are available at three filtration rates using the same suspended solids concentration and character. For example, runs F-1-I, II, III and F-2-I, II, III were made at flow rates of 2, 4 and 6 gpm/sq. ft. In the former, the C_o was 12.5 mg/ℓ; in the latter, C_o was 12.9 mg/ℓ. Thus, the C_o level was fairly equivalent and the headlosses were expected to approximate each other. Fig. 27 shows a plot of headloss developed as a function of the total volume of filtrate in each run. Surprisingly, in both runs, the headloss appears to be a function of the total volume of water filtered, relatively unaffected by filtration rate. In both runs, after 6,000 gallons of water was filtered, the headloss through the filter varied very little; i.e.:

	Headloss - ft. of water	
Rate, gpm/sq. ft.	Run F-1	Run F-2
2.0	8.8	9.1
4.0	7.2	8.6
6.0	7.9	7.6

Fig. 28 shows a plot of the headloss developed as a function of the SS accumulation within the filter pores. With the same SS concentration and characteristics, the headloss developed was dependent on the amount of SS accumulated and was not significantly affected by the flow rate.

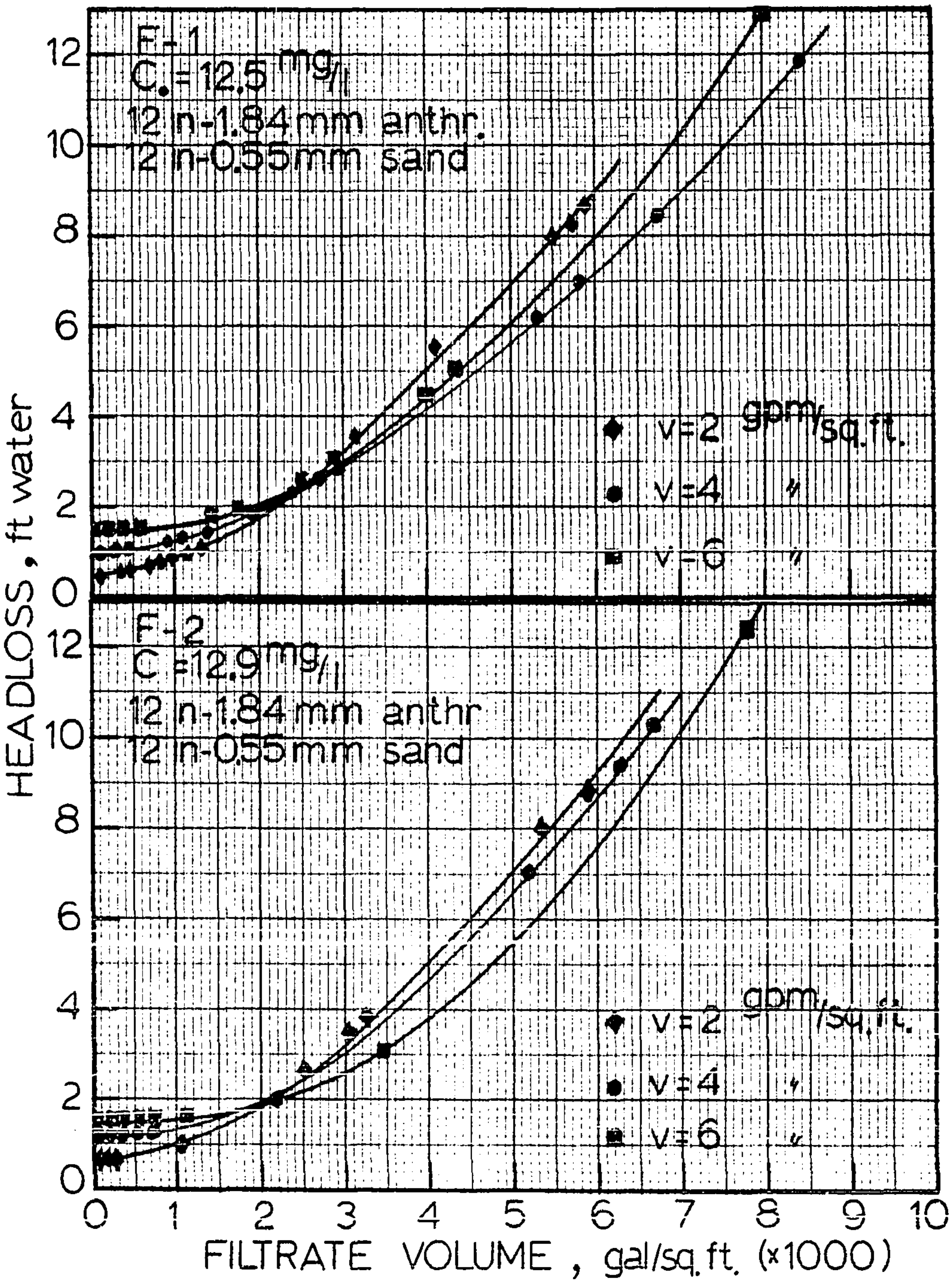


Fig. 27. Headloss vs. filtrate volume at various flow rates

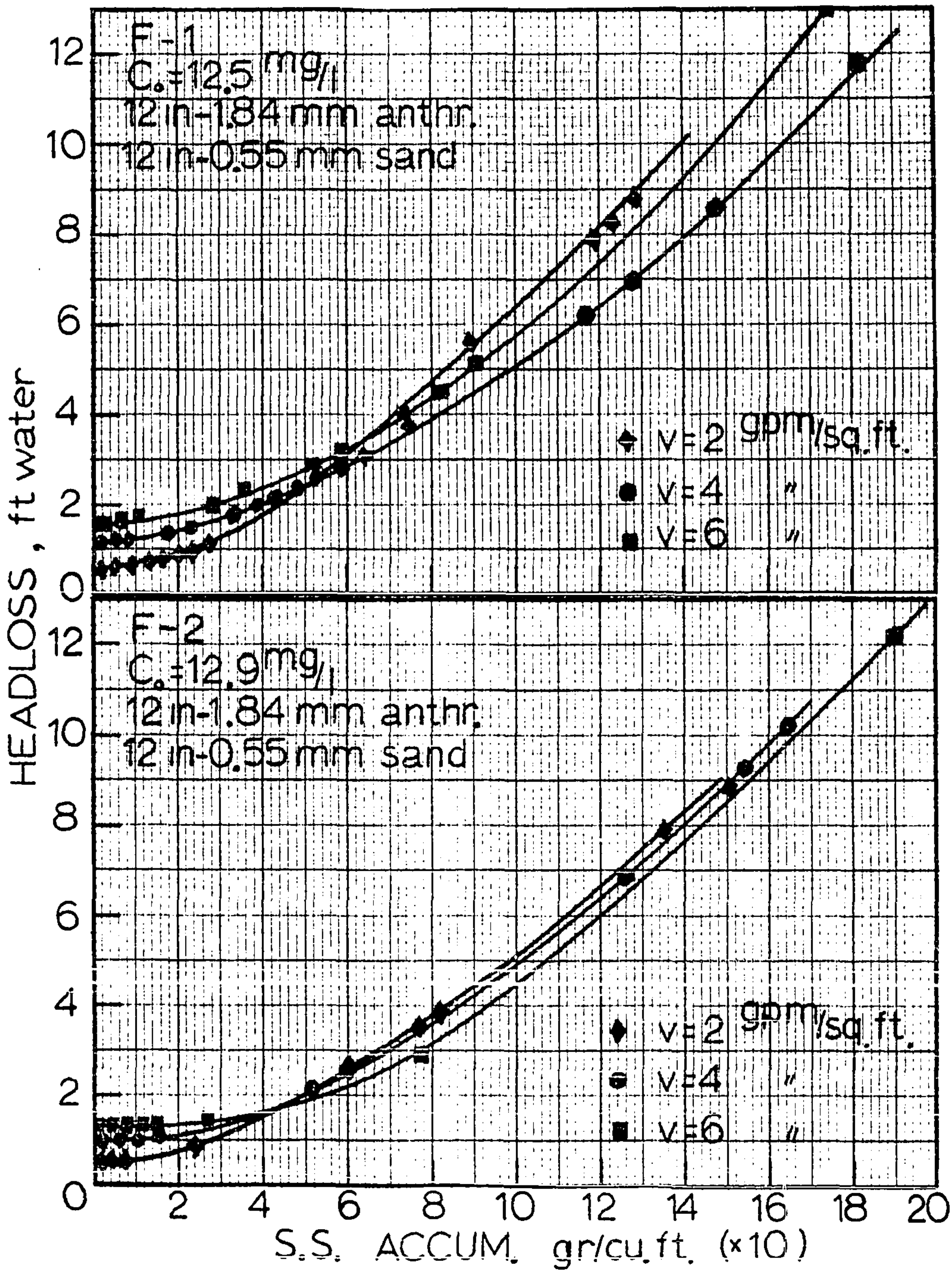


Fig. 28. Headloss vs. SS accumulation at various flow rates

This confirms the assumption made by Tchobanoglous and Eliassen (85).

The results shown in Figs. 27 and 28 indicate that filter sets which produced the same volume of filtrate accumulated the same amount of SS within their filter pores, since headloss depends on both volume of filtrate and amount of SS accumulated. This inference is in accordance with the observations made from Fig. 23, which shows that the same effluent quality was obtained from a media depth of 24 in., regardless of flow rate, when the filter sets filtered the same SS concentration and characteristics.

It should be noted that the distribution of SS accumulation in the filter bed was significantly affected by flow rate, although the total amount of SS accumulated was relatively equal at a media depth of 24 in. Fig. 29 shows the distribution of SS accumulated in various layers of the filter bed for an F-1 run, in which three filter sets were operated at various flow rates but filtered the same wastewater. The following observations can be drawn from Fig. 29:

(1) Within the top 1 in. layer, the greatest SS accumulation resulted from the lowest flow rate of 2 gpm/sq. ft., and the least SS accumulation resulted from the highest flow rate of 6 gpm/sq. ft. Rather obviously, the higher flow rates

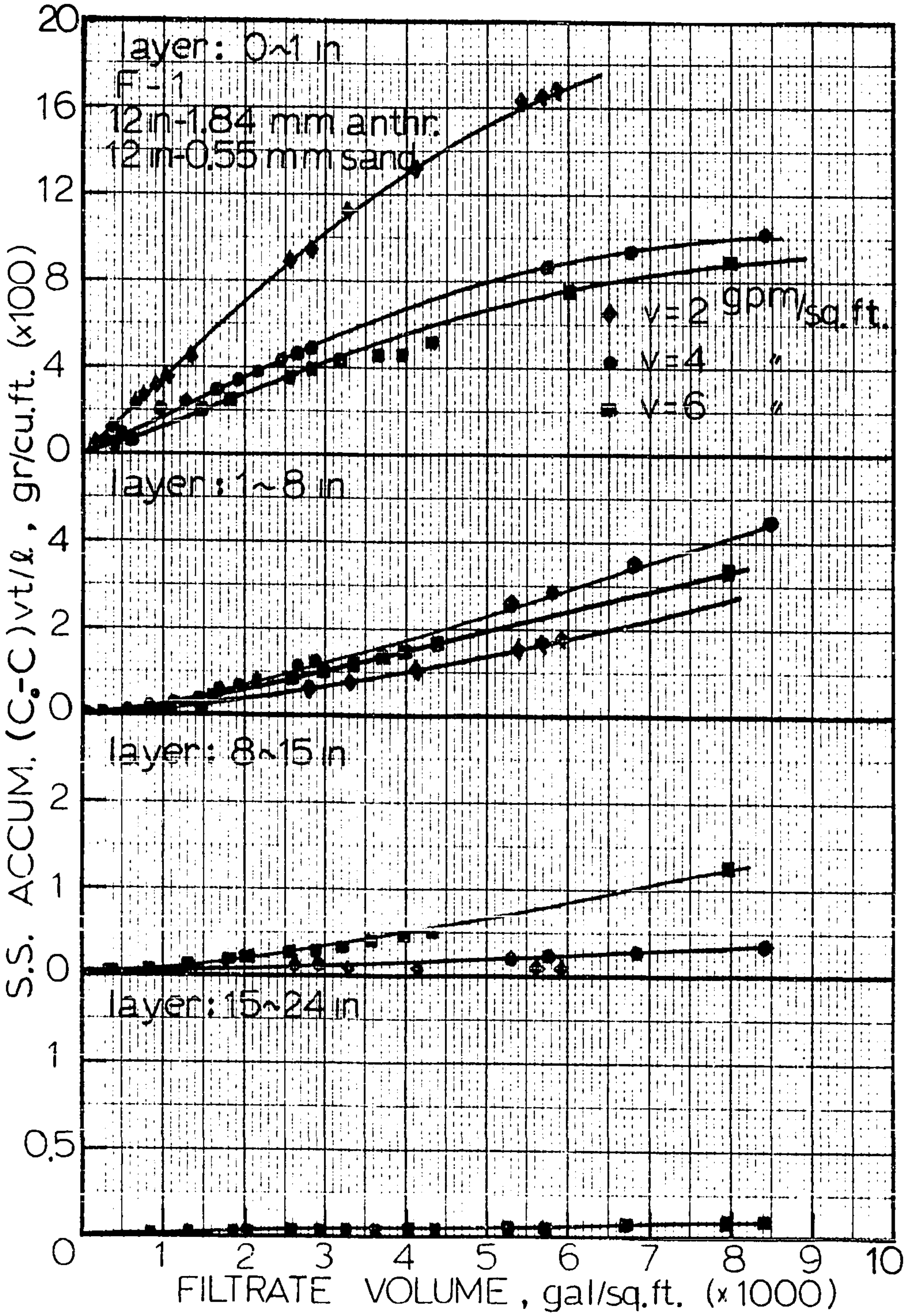


Fig. 29. SS accumulation vs. filtrate volume at various layers of filter bed

carried more of the SS deeper into the filter media.

(2) From 1 to 8 in., the greatest SS accumulation resulted from a flow rate of 4 gpm/sq. ft. Since most of the solids were removed in the top 1 in. at the 2 gpm/sq. ft. rate, there was little remaining SS to be removed in the 1-8 in. layer. At the 6 gpm/sq. ft. rate, however, the SS were carried even further into the bed.

(3) From 8 to 15 in., the greatest SS accumulation resulted from the highest flow rate of 6 gpm/sq. ft.

(4) Within the layers from 15 to 24 in., the greatest SS accumulation resulted from the highest flow rate, 6 gpm/sq. ft., followed by the results from 4 and 2 gpm/sq. ft. At these flow rates, most of the solids had already been removed in the upper layers of the media.

The distribution of SS accumulation shown in Fig. 29 indicates that a more uniform distribution of SS along the filter bed resulted from use of the higher flow rate (6 gpm/sq. ft.). For a low flow rate (2 gpm/sq. ft.), most of the SS accumulation occurred within the surface layer. This phenomenon is further verified in Fig. 30, in which the distribution of SS accumulation when the volume of filtrate

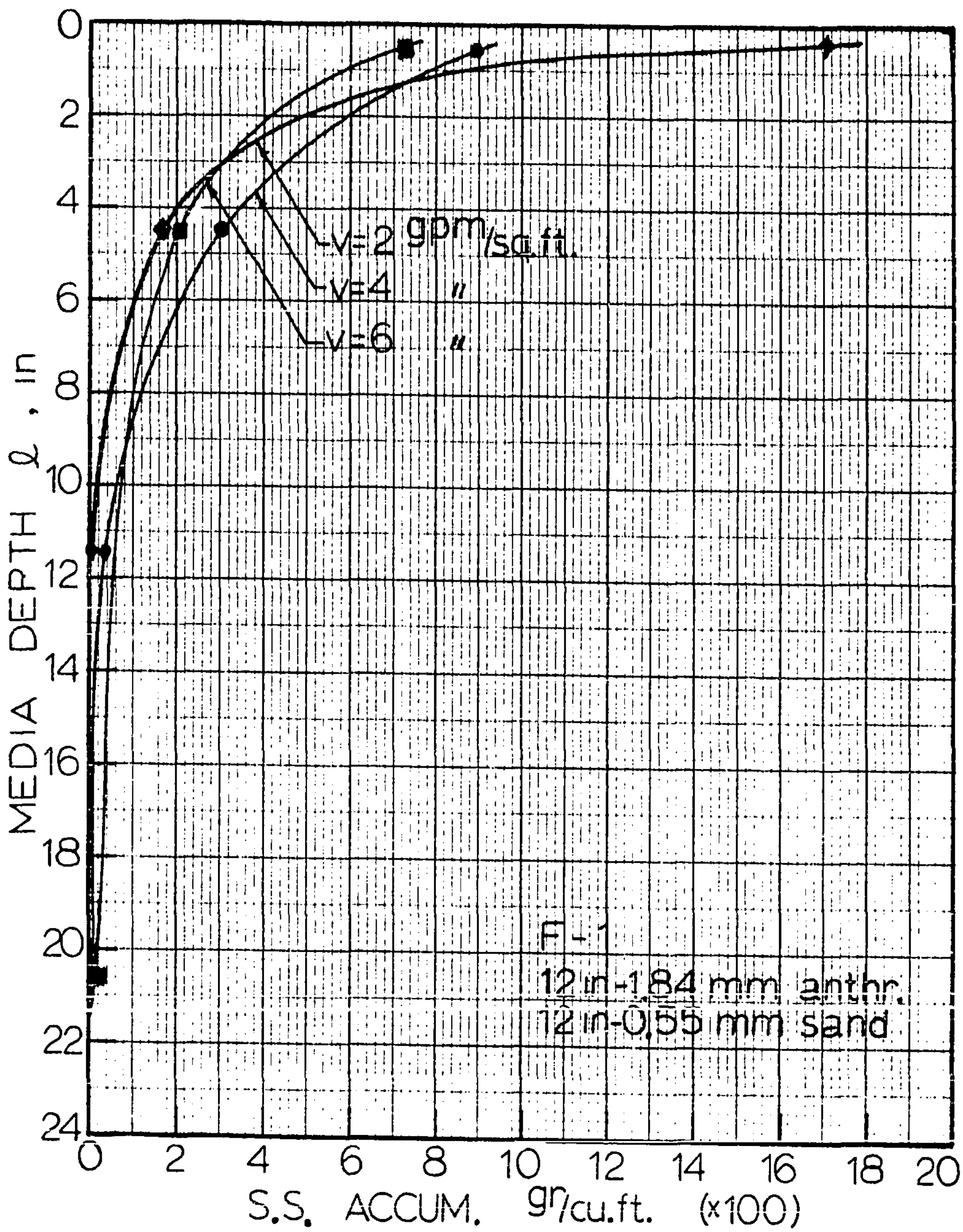


Fig. 30. SS accumulation at filtrate volume of 6,000 gal/sq. ft. vs. filter media depth

was 6,000 gal/sq. ft. was shown along the filter bed. The values shown in Fig. 30 are obtained from Fig. 29 by drawing a vertical line through the point which indicated 6,000 gal/sq. ft. filtrate volume. The results shown in Fig. 30 can be summarized as follows:

	2 gpm/sq.ft.		4 gpm/sq.ft.		6 gpm/sq.ft.	
layer 0~1 in.	1700	gr/cu.ft.	900	gr/cu.ft.	740	gr/cu.ft.
layer 1~8 in.	180	"	300	"	250	"
layer 8~15 in.	5	"	20	"	80	"
layer 15~24 in.	-		7	"	10	"

These data indicate that, with the SS in the Ames trickling filter plant effluent and the 1.84 mm coal, 0.55 mm sand, the headloss is mostly dependent on the SS accumulated in the bed. The media size gradations are such that the higher flow rates carry material into the bed and redistribute the solids loading over a greater media depth.

2. Effect of the variation of the influent SS concentration on run length

Under the operational optimum design condition, the run length is the time when both the filter removal and hydraulic capacities have been exhausted at the same time. Therefore, the determination of run length depends on the capability of determining the rate of exhaustion of filter removal and hydraulic capacities. As shown in previous sections, the difficulty in predicting the headloss development was due primarily to the variation and uncertainty in the influent solids concentration of the wastewater. In this section, the dependence of run length on the influent solids characteristics will be investigated.

Fig. 31 shows the dependence of run length, which is defined as the time when headloss reaches 10 ft. water, on the influent solids concentration at various flow rates. As shown in the figure, run length decreases exponentially as the influent SS concentration increases. In filtering the same wastewater, run length depends on the flow rate used. Therefore, run length is the result of an interaction of three variables: flow rate, C_o, and influent SS characteristics.

In view of the variation in both nature and concentration of a wastewater to a tertiary filter plant, any theoretical prediction of filter performance and optimization leads to impractical results. For a practical filter design, an intensive pilot plant study should be conducted.

B. Practical Tertiary Filter Plant Design

Dual media filtration has been practiced for some time. In design, the selection of the media sizes and combinations

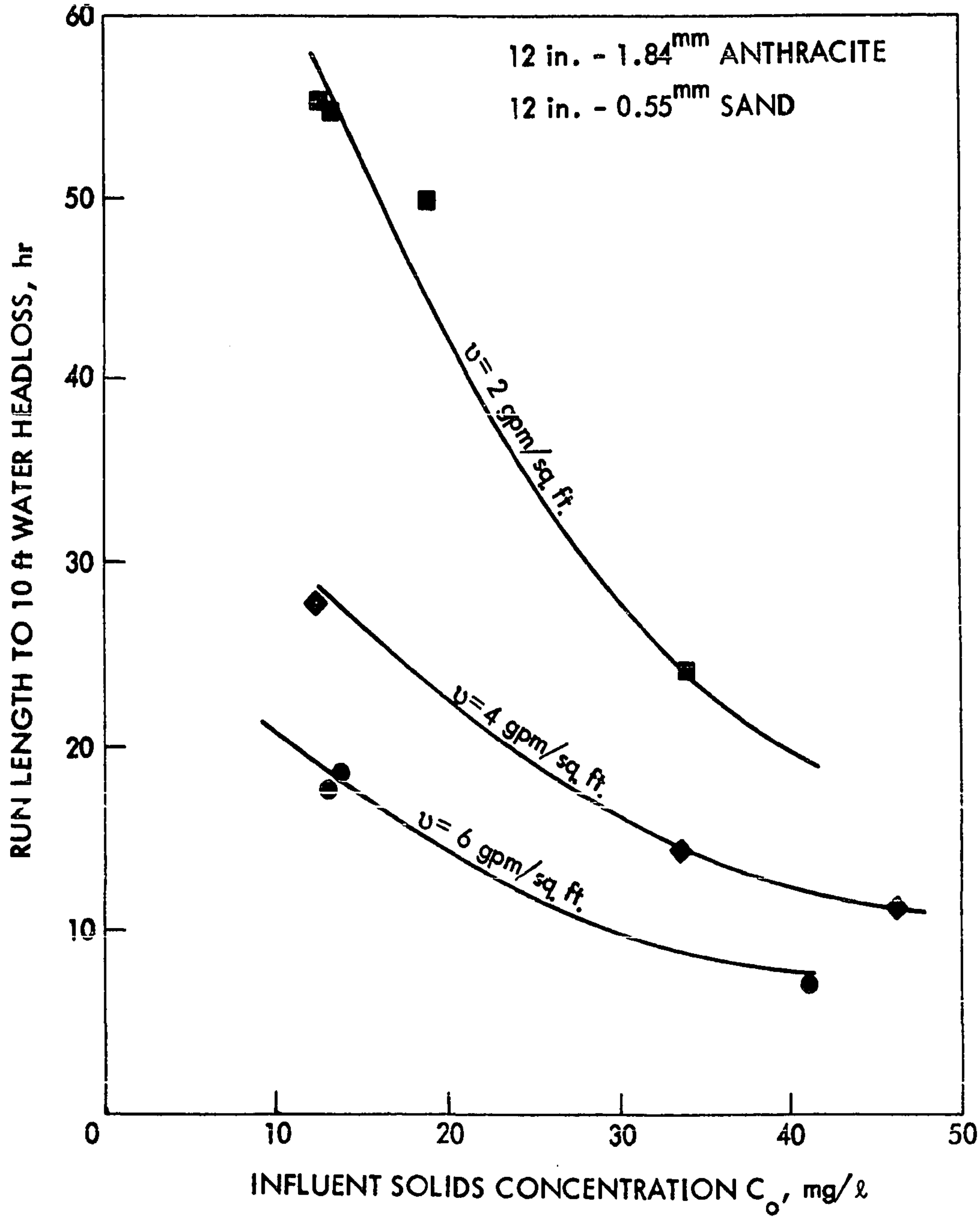


Fig. 31. Run length vs. influent SS concentration at various flow rates

as well as the respective media depths, have been based primarily on a minimum of past experience. A rational method for designing optimal dual media filters (media sizes and depths) is still not available. However, in this study, the writer proposes (and demonstrates) a rather crude but applicable method for selecting the proper anthracite-sand media size combinations and depths required based on the data from his pilot plant study (phase A through phase E).

As developed in the previous chapter (Chapter V), 12 to 15 in. of uni-sized 1.84 mm anthracite on top of 12 to 15 in. of uni-sized 0.55 mm sand in an anthracite-sand dual media filter was found to be a proper selection.

The remaining task for completing the design of a tertiary filter plant is to select a flow rate and run length to maximize water production per sq. ft. of filter. The variability and complexity of the influent SS characteristics prevents optimum selection of flow rate and run length for a tertiary filter plant. However, data collected from a pilot plant study can be used to arrive at a practical selection of flow rate and run length.

Data from Fig. 31 can be replotted showing the relationship between run length and flow rate at various influent solids concentrations, as shown in Fig. 32. From Fig. 32, it is possible to predict the run length at

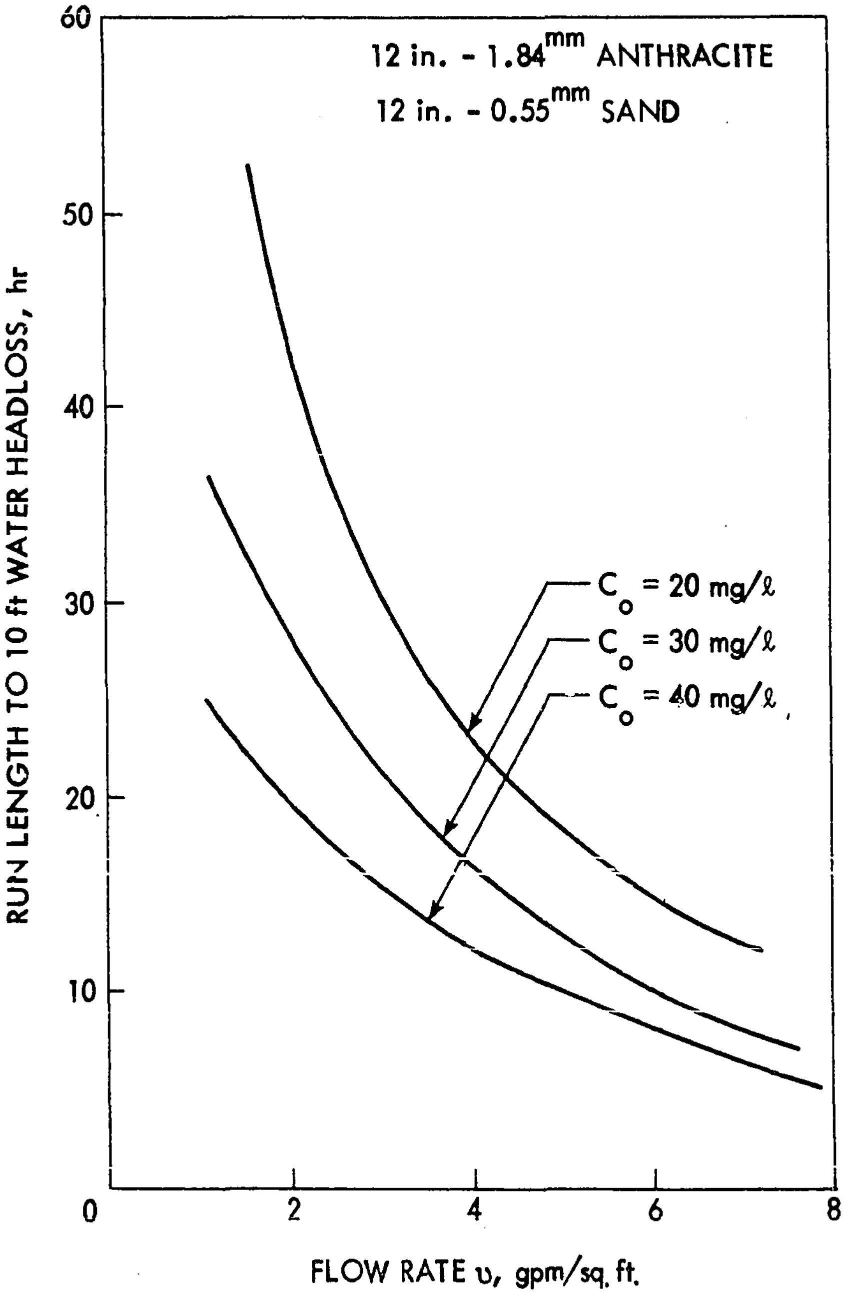


Fig. 32. Run length vs. flow rate at various influent SS concentrations

various flow rates at certain influent solids concentrations. For example, the run length to 10 ft. water will be 12.5 hr., if the flow rate is 4 gpm/sq. ft. and the influent SS concentration is 40 mg/ℓ. As indicated in Fig. 32, there are several alternative choices of flow rate to reach a certain predetermined headloss, say 10 ft. water, under a fixed range of influent solids concentration. For example, with influent solids concentrations of 40 mg/ℓ, run lengths will be 20, 12.5, and 8.5 hr. respectively, if flow rates are 2, 4, and 6 gpm/sq. ft. Similarly, with influent solids concentrations of 30 mg/ℓ, run lengths are 28.5, 16.5, and 10 hr., respectively, if flow rates are 2, 4, and 6 gpm/sq. ft.

As a result, several designs can be adopted, using higher flow rate with shorter run length, or using a lower flow rate to obtain a longer run length. An engineer who is responsible for designing a filter at least cost must first be able to determine the filters which provide equivalent performance. Two filters may be said to provide equivalent performance when they produce the same quantity and quality of filtered water from the same wastewater source during the same time period, for example one day. At this point, the engineer is in a position to determine the filter designs of equivalent performance based on the results from actual pilot plant operation.

1. Filters of equivalent performance

The amount of net water production within a period of time can be estimated, if media size combinations, flow rate and run length have been determined. A useful relationship between net water production and flow rate at various run length conditions can be developed as tabulated in Table 15. Water productions per run are computed by multiplying the flow rate by the various assumed run lengths. Run lengths ranging from a theoretically infinite length, which assumes no backwash is required, to 1 hr. in increments, are assumed. Backwash water is estimated by assuming that 5 minutes of water wash is required at a rate of 20 gpm/sq. ft., preceded by 3 minutes of air wash at a rate of 3 cfm/ sq. ft. This backwash rate expands the media bed, which is 12 in. of uni-sized 1.84 mm anthracite on top of 12 in. of uni-sized 0.55 mm sand, 20 per cent during backwash. This 20 per cent expansion criteria was recommended by Camp, *et al.* (8), and is based on their field study. Net water production per day is estimated by subtracting the backwash water required from the water produced in a run, then multiplying it by the total number of runs in a day, assuming that each filter requires 30 minutes only for a complete backwash cycle. (Appendix B includes another approach for estimating the down time required depending on the flow rate.)

Table 15. Net water productions determined by flow rates and run lengths

Flow rate gpm/sq.ft.	Run length hr.	Water prod. per run gal/sq.ft.	Backwash water* gal/sq.ft.	No. of runs** per day	Net water production gpd/sq.ft.
	∞				2,880
	50	6,000	100	0.475	2,800
	30	3,600	100	0.788	2,760
	20	2,400	100	1.170	2,690
2	10	1,200	100	2.285	2,510
	5	600	100	4.370	2,180
	3	360	100	6.85	1,780
	2	240	100	9.60	1,345
	1	120	100	16.00	320
	∞				5,760
	50	12,000	100	0.475	5,650
	30	7,200	100	0.788	5,570
	20	4,800	100	1.170	5,500
4	10	2,400	100	2.285	5,260
	5	1,200	100	4.370	4,810
	3	720	100	6.850	4,250
	2	480	100	9.60	3,640
	1	240	100	16.00	2,240

* 5 min. water backwash at rate of 20 gpm/sq. ft.
** 30 min. down time per run assumed.

Table 15. (continued)

Flow rate gpm/sq.ft.	Run length hr.	Water prod. per run gal/sq.ft.	Backwash water* gal/sq.ft.	No. of runs** per day	Net water production gpd/sq.ft.
	∞				8,640
	50	18,000	100	0.475	8,500
	30	10,800	100	0.788	8,440
	20	7,200	100	1.170	8,410
6	10	3,600	100	2.285	8,000
	5	1,800	100	4.370	7,440
	3	1,080	100	6.850	6,700
	2	720	100	9.60	5,950
	1	360	100	16.00	4,160

As shown in Table 15, there exists an upper limit of net water production at each flow rate. The maximum net water production which can be obtained in a day is 2,880, 5,760 and 8,640 gallons per square foot for flow rates of 2, 4, and 6 gpm/sq. ft., respectively.

The results from Table 15, plotted in Fig. 33, show the relationship between net water production and flow rate at various run lengths. Points of the same run length at different flow rates are connected by straight lines to show the locus of equal run length. Each horizontal line represents equal net water production line. Assume that a net water production of 3,500 gal/day/sq. ft. is desirable. This is unattainable with any flow rate less than 2.45 gpm/sq. ft. even with an infinite run length. At 2.5 gpm/sq. ft., a run length of 50 hr. is needed. At a flow rate of 3 gpm/sq. ft., a run length of 5 hr. is needed. At a flow rate of 3.9 gpm/sq. ft. a run length of only 2.0 hr. is needed. With a one hour run, desired production can be reached at a 5.3 gpm/sq. ft. flow rate.

Filters of equivalent performance based on the results from actual pilot plant operation can be found from Fig. 33. For example, in a pilot plant filter run, a run length of 20 hr. was obtained at a flow rate of 2 gpm/sq. ft., when it was filtering wastewater with an influent suspended solids concentration of 40 mg/ℓ (Fig. 32). Fig. 33 can be

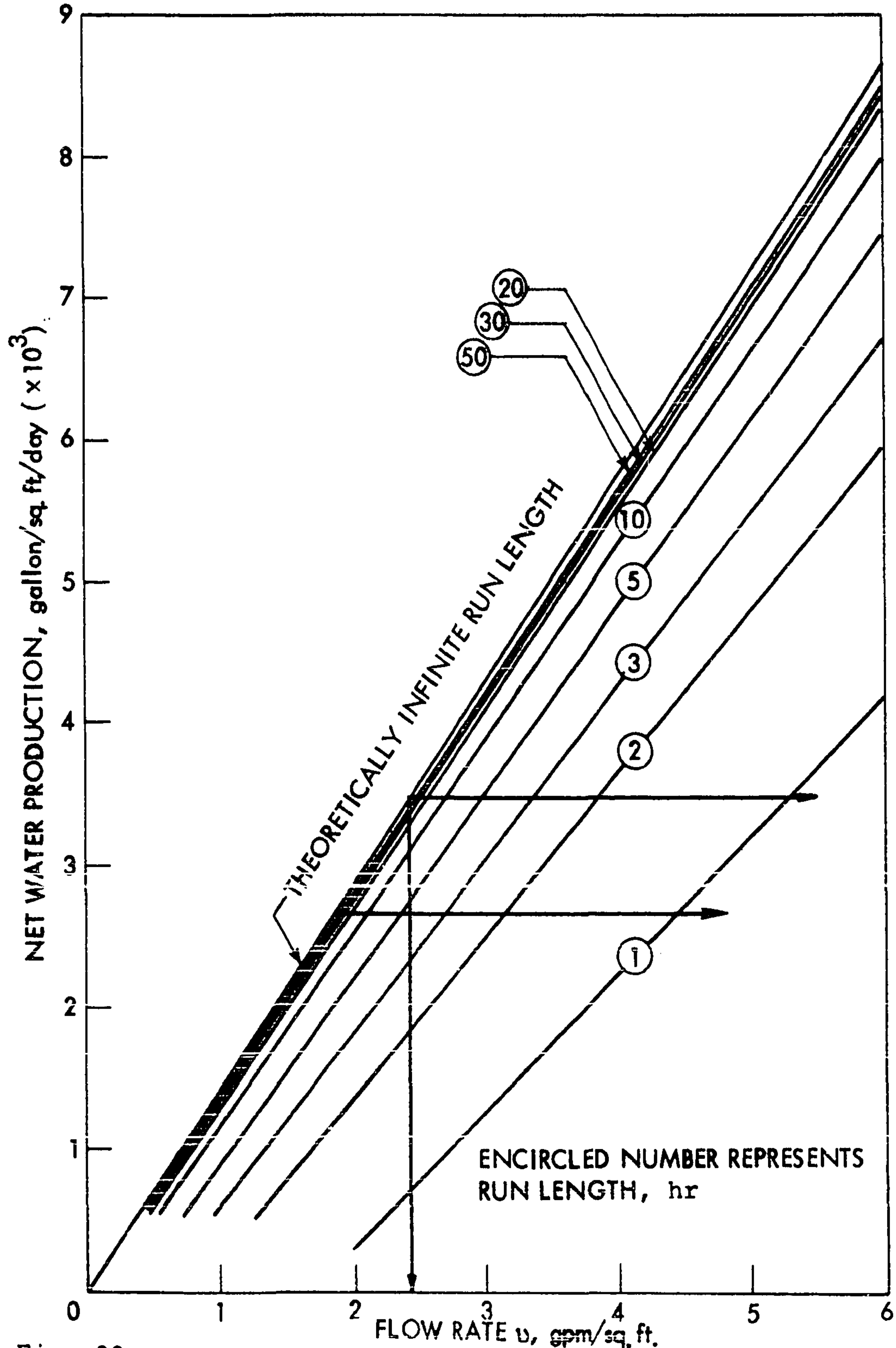


Fig. 33. Net water production vs. flow rate at various run lengths (30 min. backwashing period assumed)

used to find filters of equivalent performance by drawing a horizontal line through a point which represents this run, as shown in Fig. 33. Several designs other than that with a flow rate of 2 gpm/sq. ft. provide equivalent performance, producing 2,690 gal/sq. ft. of water per day. These filter designs of equivalent performance are listed in Table 16.

Table 16 shows that a flow rate of 3 gpm/sq. ft. with a run length of 2.5 hr. is equivalent to an actual filter run with a flow rate of 2 gpm/sq. ft. and a run length of 20 hr. since both produce 2,690 gal/sq. ft. of water per day. Any run length longer than 2.5 hr. at a flow rate of 3 gpm/sq. ft., would produce more than 2,690 gal/sq. ft. of water per day and would be a more efficient filter design. The curve for C_o = 40 mg/ℓ in Fig. 32 indicates that with a flow rate of 3 gpm/sq. ft., the run length can be expected to approximate 15 hr. This shows that a more favorable design can be obtained, if a higher flow rate is used. The results from pilot plant operation showed that a flow rate ranging from 4 to 6 gpm/sq. ft. could be successfully used with an expected run length ranging from 12.5 to 8.5 hr. respectively when the influent SS concentration of wastewater was 40 mg/ℓ.

Fig. 33 indicates that the maximum water production available with an infinite run length made at 6 gpm/sq. ft. would be 8,640 gallons. With C_o = 40 mg/ℓ, only an 8.5 hr.

Table 16. Predicted filters of equivalent performance (C_o = 40 mg/ℓ)

	Actual* run data		Predicted filters of equivalent performance, if t values are attainable								Net water production
Line	v gpm/ sq.ft.	t hr.	v gpm/ sq.ft.	t hr.	v gpm/ sq.ft.	t hr.	v gpm/ sq.ft.	t hr.	v gpm/ sq.ft.	t hr.	gal/sq.ft./ day
1	2	20	3	2.5	4	1.4	5	0.8	6	0.5	2,690
2	4	12.5	3.8	30	3.9	20	4.5	4	5	2.7	5,300
3	6	8.5	6.3	5	5.9	10	5.65	20	5.6	30	7,800

* Influent suspended solids concentration was 40 mg/ℓ (Fig. 30).

can be expected at that flow rate. Fig. 33 also suggests that the flow rate - length of run combinations in line 3 of Table 16 would also give the same net water production of 7,800 gallons per sq. ft. per day. These can be summarized as follows:

	Flow rate	Run length needed, hr.	Run length expected, hr. (Fig. 32)
	6.3	5.0	7.5
Production = 7800 gpd/sq. ft.	5.9	10.0	8.0 (not feasible)
	5.65	20.0	10.6 (not feasible)
	5.6	30.0	11.0 (not feasible)

Therefore, only a 6.0-6.3 gpm/sq. ft. would be capable of producing a minimum of 7,800 gpd/sq. ft. of filter.

Therefore, the availability of results from a limited number of filter runs in a pilot plant can generate sets of filter designs with equivalent performance through the use of Fig. 33. Thus, filtration cost can be compared among filter designs of equivalent performance, from which a practical filter design with reasonable cost can be obtained.

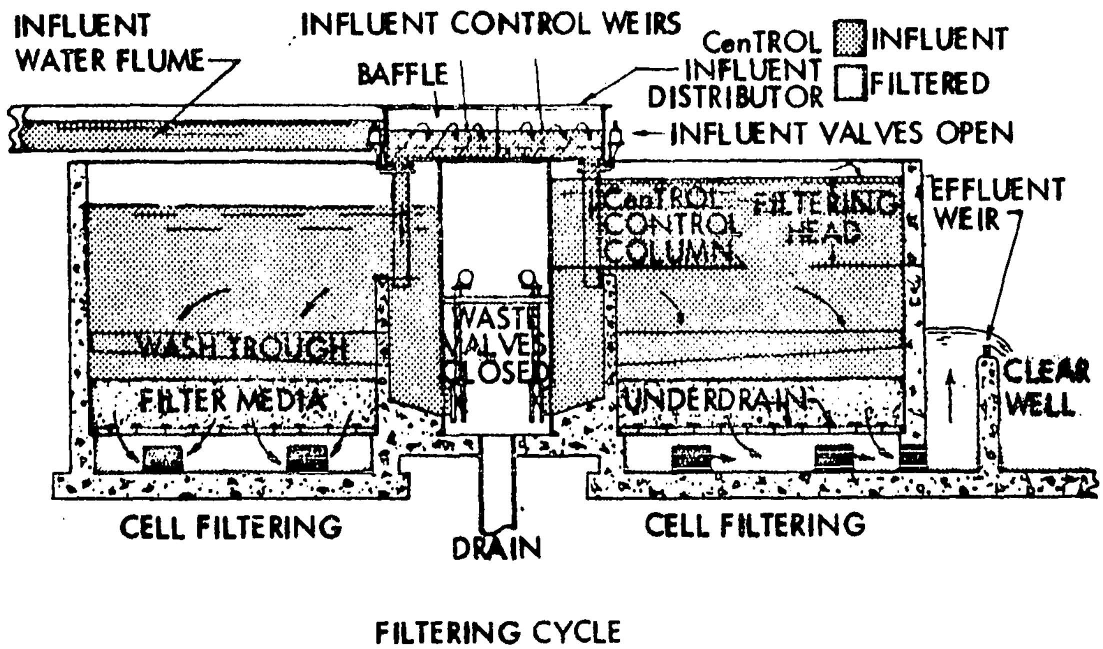
2. Filtration cost

Costs involved in tertiary filtration of wastewater are difficult to obtain. Usually, filtration cost data are buried in total general contract bidding and are not readily available. Most literature (55, 77, 94) which reported the cost of granular filters for tertiary wastewater treatment related the capital cost in dollars to the plant capacity in MGD. Usually both the operational and maintenance cost are excluded from published cost estimations. This is due to the fact that both operational and maintenance costs are rather difficult to predict, estimate, or determine.

The primary purpose of this study is not intended to provide filtration costs for general use in plant design but to evaluate the sensitivity of capital cost in dollars due to the change in flow rate for various plant capacities. Cost estimation is based on the cost information provided by General Filter Company for the CenTROL filter.[1]

a. <u>CenTROL filter</u> The filter considered is a 4-cell CenTROL gravity filter (Fig. 34) using media size and depth as determined in the previous chapter. This type of filter employs a central control principle with automatic flow distribution and without mechanical rate of flow controllers. Wastewater flows through an inlet flume into the distributor, under stilling baffles and over the filter cell inlet weirs which divide the flow equally and automatically among the filters in operation. From the filter cell inlet weirs, the wastewater flows into each filter cell

[1]Product of General Filter Company, Ames, Iowa.



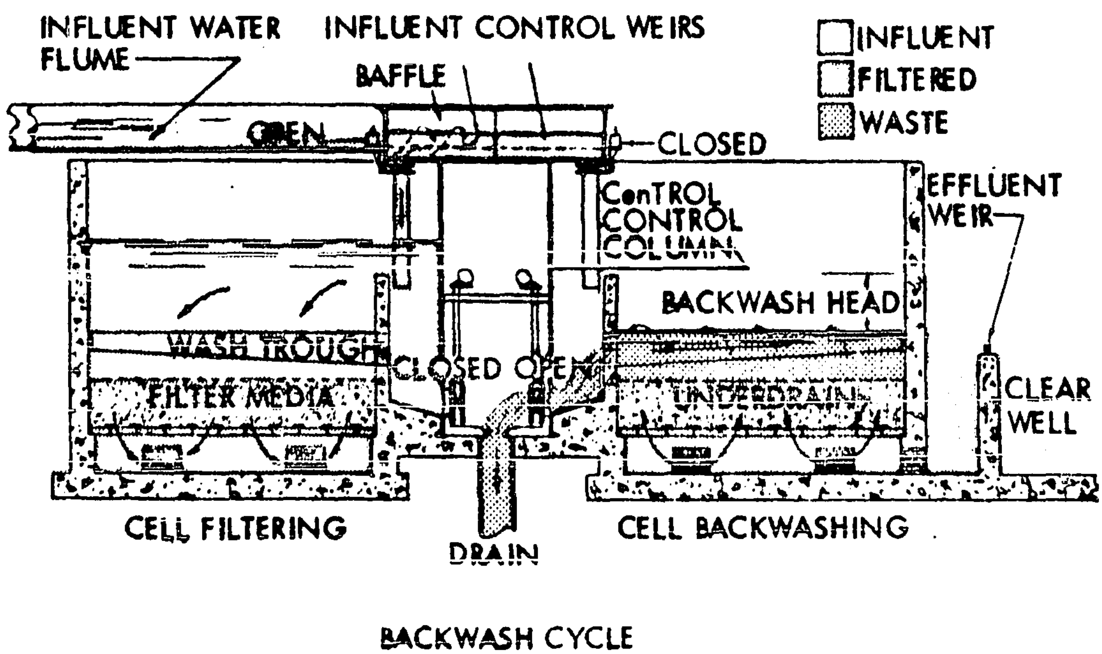


Fig. 34. Filtering and backwash cycles of a CenTROL filter

gullet through inlet lines and then through the filters.

Filtered water is collected through an underdrain chamber common to all filter cells. The filtered water is discharged over an adjustable effluent control weir to the clear well. No flow controllers are required. The outlet weir elevation is above the top of the filter media. Negative head, therefore, cannot exist in the filters. The filtering head is the difference in elevation between the effluent control weir and the water level in the filter. The maximum filtering head available is determined by the highest water level which could exist above the filter media. The height of the walls of the filter plant determines the head which is available for filtering.

The filtering head is observed by noting the operating water levels in each filter cell. As the headloss through the filter increases, the operating water level rises until it reaches the maximum permissible level above the filter near the top of the filter wall.

A 3 minute air wash with an air rate of 3 cfm/sq. ft. is followed by a 5 minute water wash with a rate of 20 gpm/ sq. ft., which is provided by a backwash pump.

The CenTROL filters are operated under constant rate conditions except during the backwash of one of the filters. When one filter cell is backwashing, the influent normally directed to it is directed uniformly to the other filters in operation. With four filter cells in the CenTROL filter battery and one backwashing, each of the other three cells would receive a rate increase of 33.3 per cent of the flow directed to it over a reasonably long period of time for rate adjustment.

b. Capital cost   Filtration cost included in this study is based on the capital cost, which includes the equipment cost and cost of filter structure.

Items involved in equipment cost estimates include:

(1) An aluminum inlet distributor (inlet valves not included).

(2) Inlet valves, flanged connection and inlet pipes (these mount on inlet distributors).

(3) Control column (including material and labor).

(4) Backwash waste valves and connections (these mount on the control column).

(5) Aluminum crosswalk.

(6) Wash troughs (fiberglass).

(7) Underdrain parts with tail pipes.

(8) Filter media.

(9) Airwash blower and valves.

(10) Pneumatically operated cell isolation valves.

(11) Pumps.

The cost of the equipment is governed principally by the surface area of the filter required. Therefore, before equipment costs are calculated, the total filter area required is determined on the basis of the plant flow capacity and operating flow rate.

The main item of variation in the cost of a CenTROL filter from place to place lies in the cost of the construction of the filter plant and assembly of the equipment. The contractor bids a general contract using different unit prices for concrete in place depending on plant location. However, a unit price of $200 per cubic yard was assumed for estimating the concrete cost. Structural costs required in the construction cost estimate are:

(1) Cost of filter floor.

(2) Cost of underdrain slab.

(3) Cost of effluent box.

(4) Cost of filter wall and cell wall.

The profit earned by the general contractor is assumed to be 10 per cent of the total cost of the equipment provided by the manufacturer. The legal and engineering fees involved are assumed to be 20 per cent of the total cost of the equipment and concrete provided by the general contractor.

Table 17 includes the estimated capital cost based on the cost information of CenTROL filters. Three plant flow capacities are studied. In each case, the filter area required, equipment cost and structural cost are determined for flow rates of 2, 4, and 6 gpm/sq. ft. The relationship

Table 17. Capital cost of CenTROL filters

Plant capacity MGD	Flow rate gpm/sq.ft.	Filter[a] area sq.ft.	Equipment[b] cost $(x10^3)	Structural[b] cost $(x10^3)	Sub-total $(x10^3)	Capital[c] cost $(x10^3)
1	2	462	46	25.8	71.8	86.2
	4	231	30.3	12.2	42.5	50.5
	6	154	27.5	6.7	34.2	40.8
5	2	2,310	120	134	254	303
	4	1,155	66.6	61	127.6	153
	6	770	55.7	44.5	100.2	123
10	2	4,620	220	268	488	586
	4	2,310	110	122	232	278
	6	1,540	88.4	89	177.4	213

[a]33 per cent increase to compensate for backwash.

[b]Cost of general contractor to engineer.

[c]20 per cent of subtotal as legal and engineering fees.

between capital cost and the plant flow capacity is shown in Fig. 35. A published capital cost from Weber, et al. (94) is included for comparison.

For a certain plant flow capacity, for example 10 MGD, capital cost relating to the flow rate can be replotted, as shown in Fig. 36, to show the sensitivity of the capital cost due to the change in flow rate. As shown in the figure, capital cost decreases as flow rate increases for the same flow capacity. This indicates some saving in capital cost can be obtained, if a higher flow rate is used. Capital costs can be expected to be $280,000 and $210,000 if flow rates are 4 and 6 gpm/sq. ft., respectively, for a plant flow capacity of 10 MGD.

As a result, it appears that a higher net water production within a fixed period of time and saving in capital cost can be obtained, if the higher flow rates are used. Results from pilot plant operation showed that filters could be operated successfully at a flow rate as high as 6 gpm/sq. ft. with a reasonable run length (8.5 hr.) when the influent solids concentration of the wastewater was as high as 40 mg/ℓ.

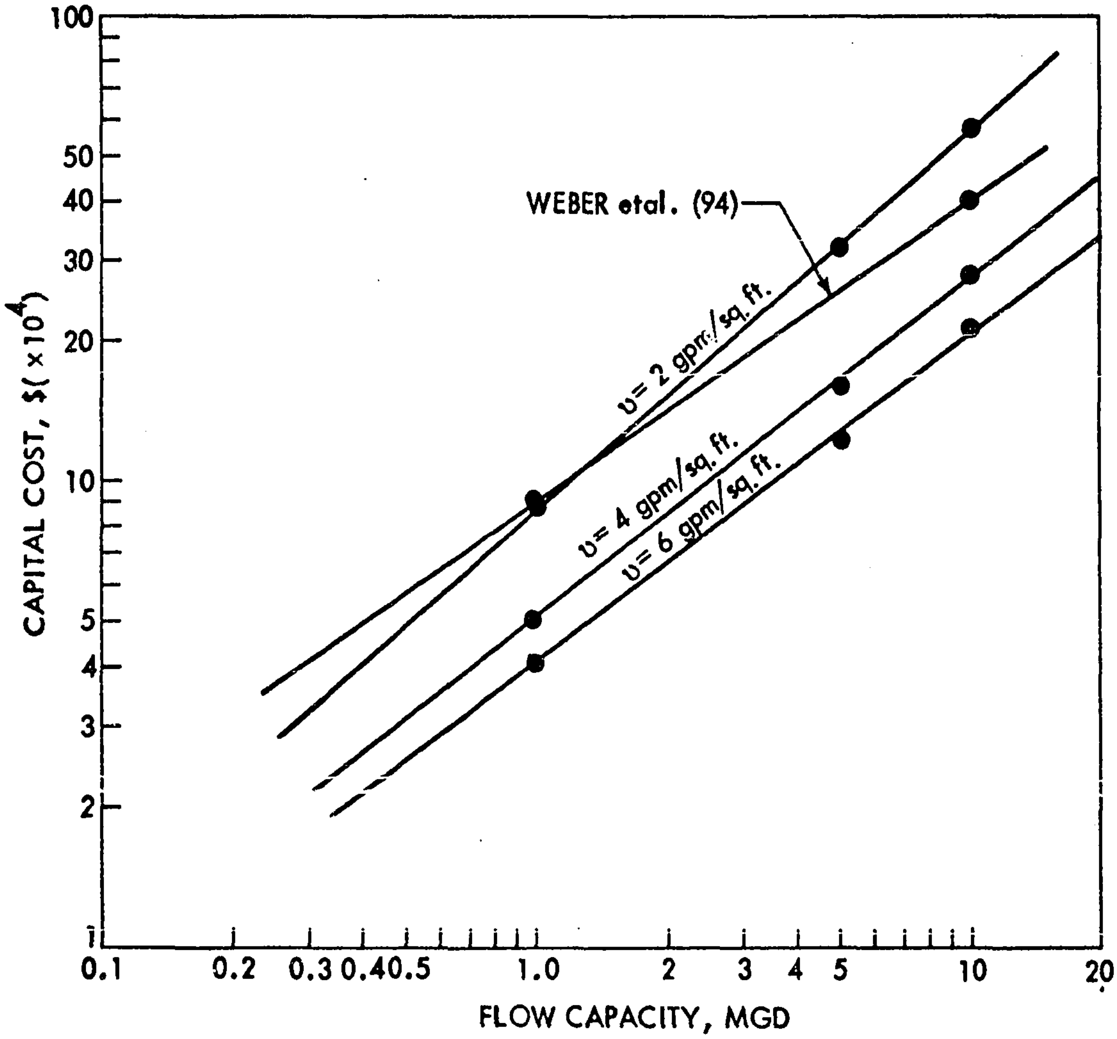


Fig. 35. Capital cost vs. designed flow capacity for various flow rates

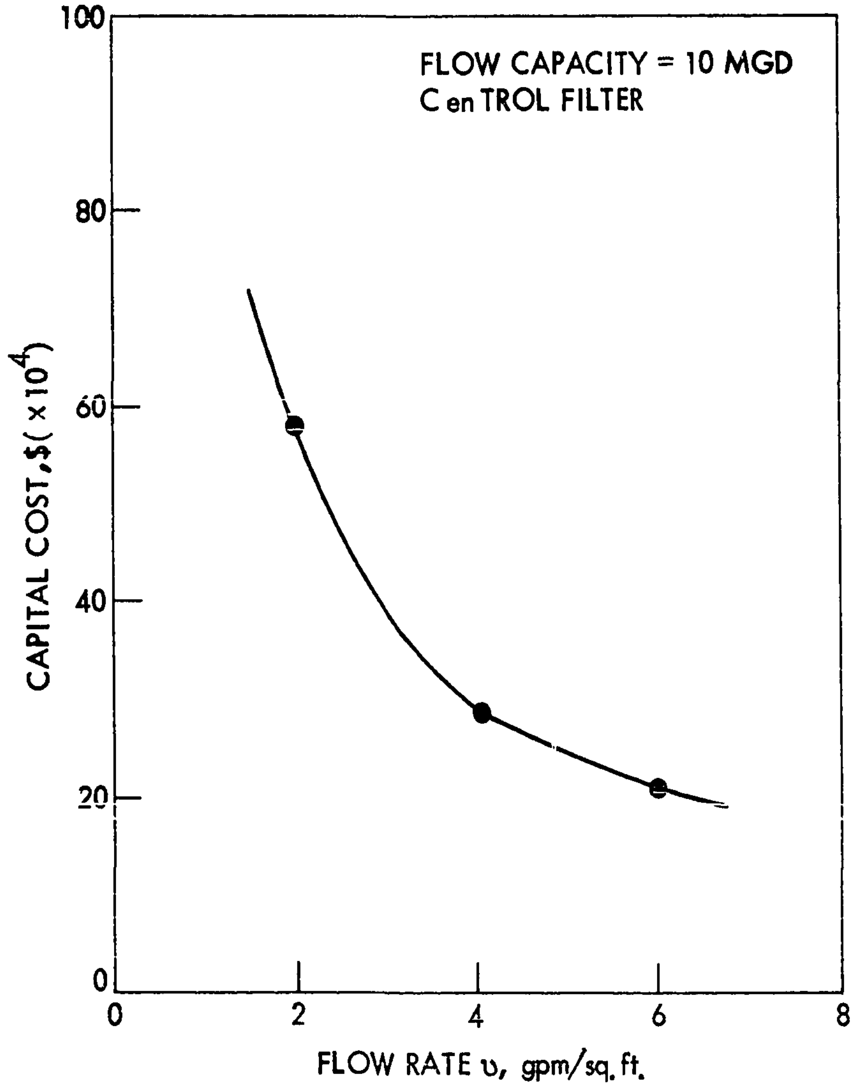


Fig. 36. Capital cost vs. filtration rate for a designed flow capacity of 10 MGD

C. Case Study - Ames Tertiary Treatment Filter Plant

The design of a wastewater treatment plant is based principally on two major factors - the characteristics and flow rate of the wastewater. Not only are the typical characteristics and the average plant flow rates important, but the hourly, daily and monthly variations in these parameters result in significant variations in operational requirements and plant performance.

1. Wastewater characteristics

The wastewater characteristics of the Ames pollution control plant were described in detail in Chapter IV. In short, the wastewater of the Ames plant consists primarily of domestic waste with a small portion of industrial waste. The wastewater characteristic which is of interest to the design of a tertiary filter plant is that of the secondary final effluent, of which the concentrations of suspended solids are of primary concern. It has been found that the plant effluent suspended solids were approximately 20-40 mg/ℓ during cold weather and 10-20 mg/ℓ during warm weather. The BOD in the final effluent ranges from 30 to 60 mg/ℓ in winter and from 15 to 25 mg/ℓ in summer.

In Ames, there is a fairly typical variation in the characteristics of the final effluent during a day, the highest strength of final effluent occurring at the noon hour, and the lowest strength occurring about 8 a.m. A typical daily variation in turbidity, SS and BOD was shown in Fig. 1, in which the highest and lowest turbidities were 19 and 7.6 JTU, respectively; the highest and lowest SS were 29 and 16 mg/ℓ, respectively; and the highest and lowest BOD were 40 and 11.5 mg/ℓ, respectively.

2. Wastewater flow

The wastewater entering the treatment plant fluctuates in flow rate over monthly, weekly, daily and hourly cycles and from dry to wet years. According to a special report[1], the monthly flow variation cycle is such that the flow rates during the winter months and in the spring are about 90% and 121%, respectively, of the yearly average rate. The maximum daily flow can generally be expected on Mondays and hourly flow varies within any one day with maximum hourly flow occurring from 11 a.m. to noon. A 4-hr. peak flow was reported to be 144 per cent of the average daily flow.

In general, there are two approaches in handling the variations in the flow rate in designing a tertiary treatment filter plant. They are: 1) provide flow equalizing basin in front of the filter plant and 2) provide sufficient

[1]Young, J. C., Baumann, E. R. and Kelman, S. Design considerations for expansion of the water pollution control plant, Ames, Iowa. Private communication. 1969.

filter area to meet the peak flow conditions.

a. Flow equalization approach   Flow equalization, although commonly used in industrial waste treatment practice, has been generally neglected in traditional municipal wastewater treatment facilities. There are two major objectives in the design of flow equalization basins. The first of these is simply to dampen the diurnal flow variation that normally exists in typical municipal wastewater collection systems. This is done to achieve a constant or nearly constant flow rate through the subsequent treatment processes. In this type of system, little consideration is given to controlling the quality changes that take place during storage. The major design factors are to supply sufficient air to keep the basin aerobic and to provide adequate turbulence to prevent solids deposition. The second objective of flow equalization is to provide the capacity to distribute shock loads over a reasonable period of time.

The justification for flow equalizing facilities is economic. The economic justification involves the relative costs of providing equalization facilities as opposed to increasing design capacities of the filter plant.

b. Peak flow design approach   The maximum hourly wastewater flow would be expected to occur on the maximum day of the month of highest flow during a wet year. Normally, it is not necessary to design for the single largest hourly flow rate possible since this flow rate may occur for only a few hours in any one year. In the normal operation of a treatment plant, the peak flow covers several hours and can be smoothed out by using the incoming sewer line and pumping station storage volume for flow equalization. Consequently, the peak flow rate for a 4-hr. period should be satisfactory for peak hydraulic load design purposes. In the following section, the projected peak wastewater flow rates for the Ames water pollution control plant will be used as a case study in designing a tertiary filtration plant. It is assumed to be designed for the year 1985. The projected peak flows are as follows[1]:

Year	Annual average daily flow rate, MGD	Peak 4-hr. flow rate, MGD Dry year	normal year	Wet year
1985	8.82	14.70	16.30	18.10

In this case study, the designed plant flow is assumed to be 18.1 MGD. It should be noted that the peak 4-hr. flow rate might not be experienced until the design life of the facility is approached.

[1]Young, et al. Design considerations for expansion of the water pollution control plant, Ames, Iowa, p. 186, herein.

3. Type and characteristics of filter

The engineering staff of the General Filter Company cooperated with Iowa State University personnel in the development of a series of graphs which makes a detailed analysis of equipment and structural cost possible. Thus, a CenTROL filter is considered in this case study. The filter characteristics are based on the results of this pilot plant study at the Ames pollution control plant. They are:

	Uni-sized media mm	Depth in.
Anthracite (top media)	1.84	12 to 15
Sand (bottom media)	0.55	12 to 15

With this media size combination, an intermixing zone of 6 in. is to be expected after backwashing.

4. Determination of filter area

In general, the filter area required is based on the plant flow capacity and the filtration rate. For designing a 4-cell CenTROL filter, an increase in filter area of 33.3 per cent can be adopted in order to maintain the designed filtration rate when one of the filter cells is down for repair. By using this practice, a safety factor of 1.33 is included in the design.

The design filtration rate should be determined by pilot plant operation under the worst SS concentrations of the incoming wastewater. Due to the fluctuation in the incoming flow rate, the actual filtration rate will fluctuate accordingly. Since the filter area to be provided is based on the 4-hr. peak flow, most of the time the actual filtration rate is less than the design rate, and the design filtration rate can be set at a high rate.

As shown previously in Fig. 32, run length to a predetermined headloss varies when the SS concentration of the incoming wastewater varies for the same filtration rate condition. For each filtration rate and run length, which are determined by the SS concentration in the wastewater, there is a fixed amount of net water production which can be obtained, as shown by the relationship in Fig. 33. Also, the run length will determine the portion of time when all four filter cells will be in operation. Case A in Fig. 37 shows the sequence of operation of each filter cell when the run length is only 1 hr. During each filter cycle, the following observations can be drawn:

(1) There is no chance that all four filter cells can be in operation simultaneously;

(2) at the most three filter cells can be in operation simultaneously for only two thirds of the time; and

(3) in general only two filter cells are in operation

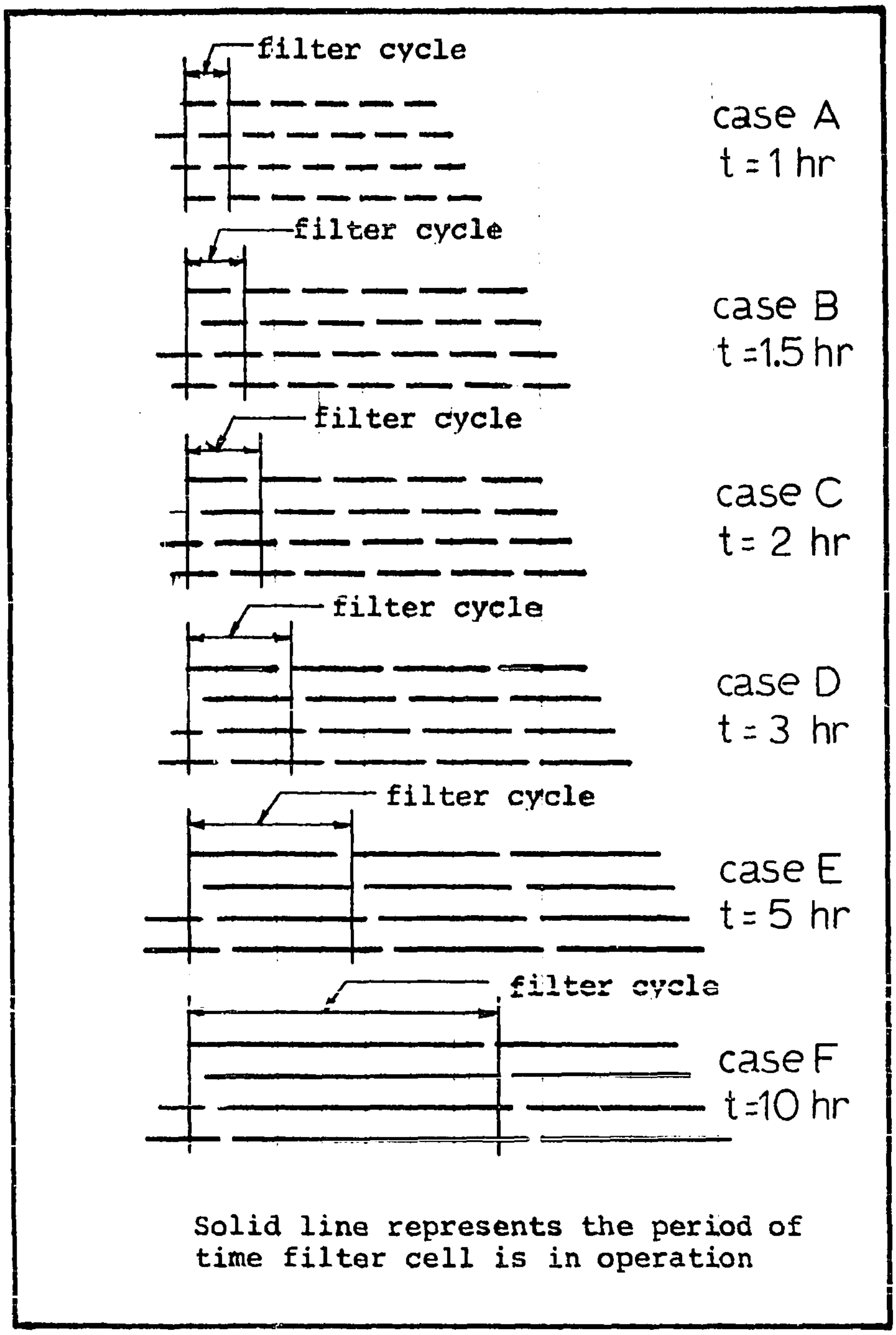


Fig. 37. Operating sequence of four CenTROL filter cells at various run length conditions

simultaneously for one third of the time.

Therefore, the filter area required should be twice that determined by assuming 4 filter cells are in operation simultaneously, to compensate for the period when only two of the four filter cells are in operation. As a result, during one third of the time when only two filter cells are in operation, the actual flow rate is the same as the design flow rate; during two thirds of the time when three filter cells are in operation, the actual flow rate is two thirds of the design flow rate.

Case B of Fig. 37 shows the condition when run length is 1.5 hr. Several observations can be drawn, as follows:

(1) there is no chance that all four filter cells can be in operation simultaneously;

(2) however, three filter cells can be in operation all the time.

Cases C, D, E, and F show the sequence of operation of each filter cell, when run lengths are 2, 3, 5, and 10 hrs., respectively. During each filter cycle, the per cent of time that all four filter cells are in operation can be summarized (Table 18 and Fig. 38).

It appears that a run length of 1.5 hr. is the critical run length for 4-cell CenTROL filters. Under the condition of a run length equal to or less than 1.5 hr., there is no chance that all four filter cells can be in operation at

Table 18. Per cent of time with four filter cells in operation related to filter run length

Cases	Run length hr.	Time when four filter cells in operation %
A	1	0
B	1.5	0
C	2	20.0
D	3	41.9
E	5	63.6
F	10	81.0

the same time. As a result, incoming wastewater will be directed to only the filter cells which are in operation. Thus, the actual filtration rate in the operating filter cells will be increased accordingly. The per cent of time that all four filter cells are in operation increases as the run length increases, as shown in Fig. 38. The per cent of increase in filtration rate in the filter in operation decreases as the per cent of time that four filters are in operation increases. This reveals the fact that run length does affect the filtration rate through the operating battery of filters.

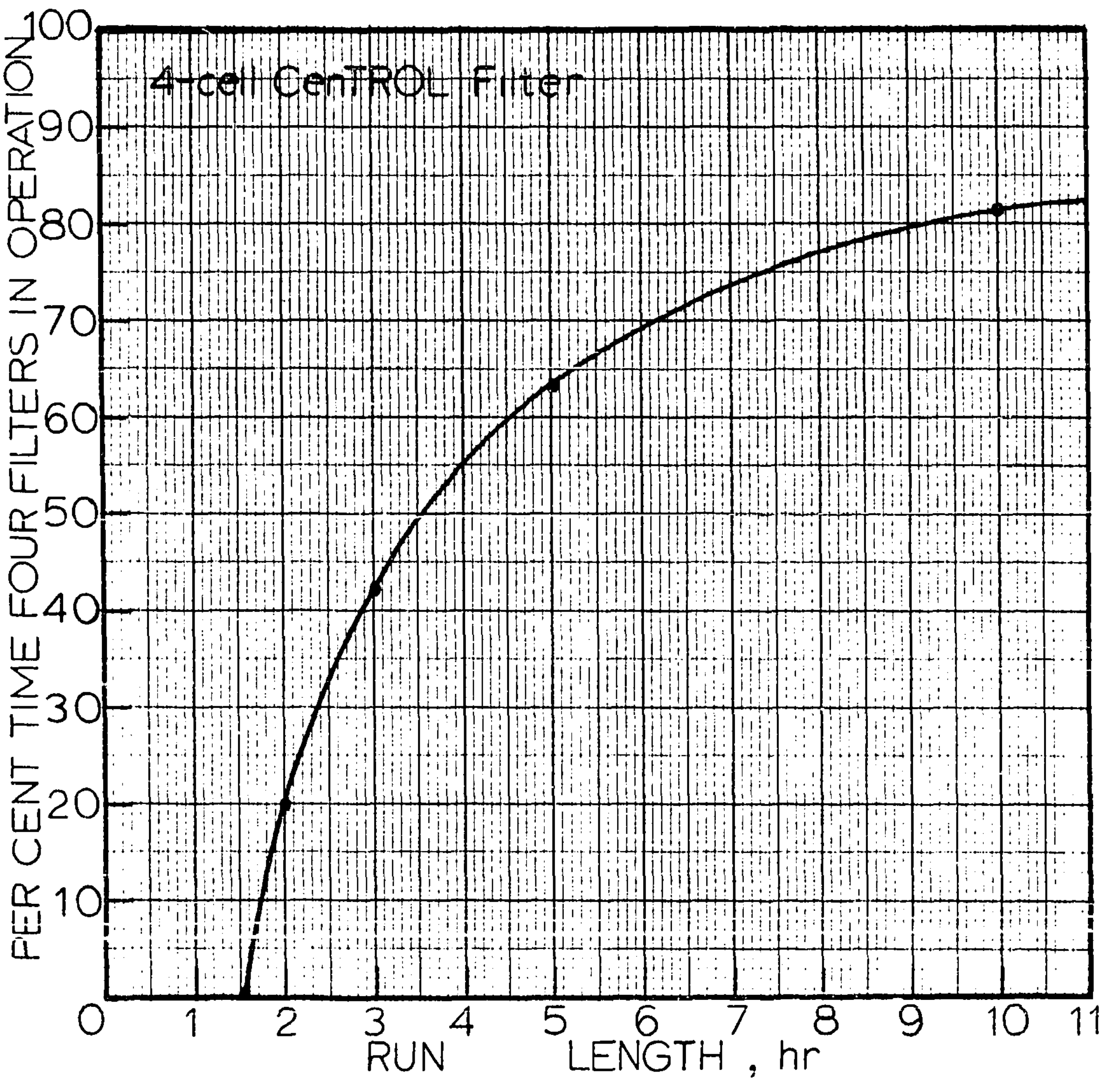


Fig. 38. Per cent of time four filter cells in operation vs. run length

5. <u>Design example 1: using one 4-cell gravity CenTROL filter</u>

a. Essential information required:

i. Projected wastewater flow rate during plant design life to 1985.

	Flow rate[1] MGD
Max. 4-hr. peak flow (wet year)	18.1
Max. daily flow (normal year)	11.3
Mean annual flow	8.82
Min. 4-hr. flow in low flow month	3.82

ii. Results from pilot plant operation:

(a) Filter media type, size and depth

Type of media	Size, mm	Depth, in.
anthracite (top media)	1.84 (10/12)	12 to 15
sand (bottom media)	0.55 (30/35)	12 to 15

(b) Backwash requirements:

3 min. air scouring at a rate of 3-5 scfm/sq. ft., followed by 5 min. water wash at a rate of 20 gpm/sq. ft., which expands the filter media selected 20 per cent during backwashing (Fig. 41, Appendix C).

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[1] Young, et al. Design considerations for expansion of the water pollution control plant, Ames, Iowa, p. 186, herein.

This degree of expansion was recommended by Camp, et al. (8).

(c) Curves relating run length to filtration rate at various influent SS concentrations (Fig. 32).

(d) Curves relating net water production from the filters to filtration rate under various run length conditions. (Fig. 33).

b. Steps of calculation

i. Determination of nominal design filtration rate

The first step in the calculation is to determine the nominal design filtration rate based on the maximum 4-hr. peak flow rate at the worst SS concentration in the wastewater expected to occur during the design life. Table 19 tabulates such a calculation assuming the maximum 4-hr. peak flow to be 18.1 MGD and the worst SS concentration in the wastewater to be 40 mg/ℓ. A portion of the filtrate is used for backwashing, which will be recycled to the final clarifier for further settling. Streander (83) reported that suspended solids accumulated in the backwash water could be settled out easily in the final clarifier. Due to the fact that the stream of water used for backwashing will reach the filter again at some later time, the filter capacity should be designed based on the maximum 4-hr. peak flow plus an estimated amount of backwashing water. Ten

Table 19. Determination of nominal design filtration rate based on max. 4-hr. peak flow (1985) at worst SS concentration
Max. 4-hr. peak = 18.1 MGD, 10% backwash water = 1.81 MGD, Co = 40 mg/l

Alt.	Flow rate, MGD	Filter capacity, MGD	Nominal filter rate gpm/sq.ft.	Nominal[a] run length hr.	Nominal filter area sq.ft.	Actual[b] filter area sq.ft.	Actual filter rate gpm/sq.ft.	Actual[a] run length hr.	Net water prod. gpd/sq.ft.[c]	Net water prod. MGD	Remarks
1	18.1	19.91	7.0	6.5	1980	2640	5.25* 7.0 **	9.0* 6.5**	6,900* 8,950**	18.2* >18.1 17.7**<18.1	Suff. Not suff.
2	18.1	19.91	6.5	7.5	2125	2830	4.89* 6.5 **	10* 7.5**	6,500* 8,500**	18.4* >18.1 18.05**<18.1	Suff. Not suff.
3	18.1	19.91	6.25	8.0	2210	2940	4.7* 6.25**	10.5* 8.0**	6,250* 8,250**	18.4* >18.1 18.2**>18.1	Suff. Suff.
4	18.1	19.91	6.0	8.5	2310	3070	4.6* 6.0**	10.6* 8.5**	6,100* 7,900**	18.7* >18.1 18.3**>18.1	Suff. Suff.
5	18.1	19.91	5.0	9.8	2760	3680	3.76* 5.0 **	12.7* 9.8**	5,050* 6,590**	18.6* >18.1 18.4**>18.1	Suff. Suff.

* from filter cells in operation.
** one of the filter cells down for repairs.

[a]From Fig. 32.

[b]1.33 times nominal filter area.

[c]From Fig. 33.

per cent of the flow is assumed to be required for backwashing in this example and this water will increase the amount of wastewater that will have to be filtered.

The determination of the nominal filter design rate proceeds by a trial and error method. As shown in Table 19, the nominal maximum flow design rate was assumed to be 7.0 gpm/sq. ft. in the first alternative. Thus, the nominal filter area required was determined as follows:

$$\text{Nominal filter area} = \frac{\text{Filter capacity required}}{\text{Assumed nominal filter rate}}$$

$$= \frac{19.91 \text{ MGD} \times 694 \text{ gpm/MGD}}{7.0 \text{ gpm/sq. ft.}}$$

$$= 1980 \text{ sq. ft.}$$

A safety factor of 1.33 was assumed to meet the condition in which one of the filter cells is down for repairs. Thus, for normal conditions, the actual filter area, 2640 sq. ft., is 1.33 times the nominal filter area. Accordingly, the actual filtration rate would be 5.25 gpm/sq. ft. (7.0/1.33) instead of 7.0 gpm/sq. ft. However, a filtration rate of 7.0 gpm/sq. ft. will be the actual filtration rate when one of the filter cells is down for repairs or for backwashing. Actual run lengths can be determined using the actual filtration rate and SS concentration in the wastewater from the curves generated from the results of the pilot plant operation (Fig. 32). From Fig. 32, the actual run lengths would be 9.0 and 6.5 hrs. when the actual filtration rates are 5.25 and 7.0 gpm/sq. ft., respectively, when C_o = 40 mg/ℓ SS. After the actual filtration rate and run length are determined, net water production under either condition (four filter cells in operation or one cell down for repairs or backwashing) can be estimated from Fig. 33. Thus, (from Fig. 33):

Filter operating condition	Net water production gpd/sq.ft.	MGD	Remarks
Four filter cells in operation	6,900	18.2	sufficient, >18.1
Three filter cells in operation	8,950	17.7	insufficient, <18.1

This result indicates that, under the assumed nominal filtration rate of 7.0 gpm/sq. ft., the filter design is sufficient to meet the maximum 4-hr. peak flow operating condition when all four filter cells are in operation. However, it is not sufficient to meet the maximum 4-hr. peak flow if one of the filter cells is down for repairs or for backwashing. Thus, it is concluded that the assumed nominal filtration rate of 7 gpm/sq. ft. is too high. Another attempt is required.

Alternative 2, as shown in Table 19, shows the second try assuming a nominal filtration rate of 6.5 gpm/sq. ft.

The result is similar to the first alternative, i.e., the filter design is sufficient to meet the maximum 4-hr. peak flow when all four filter cells are in operation but is not sufficient when one of the filter cells is out of service.

The third alternative, with an assumed nominal filtration rate of 6.25 gpm/sq. ft., shows that the filter design is sufficient to meet the flow as designed for both conditions. So do the fourth and fifth alternatives. Consequently, the highest nominal filtration which provides a filter design sufficient to meet both conditions is between 6.25 and 6.5 gpm/sq. ft. Theoretically, a further try can be made to determine the highest nominal filtration rate which can be used for this plant. However, for practical purposes, it appears that a nominal filtration rate of 6.25 gpm/sq. ft. is close enough to the highest nominal filtration rate. A filter design, which has been based on the maximum 4-hr. peak flow condition, has to be checked to see whether it is sufficient to produce filtered wastewater under all other flow conditions, such as maximum daily flow, mean annual and minimum 4-hr. in the low flow month.

ii. Meeting other flow conditions

A check to determine whether the filter can operate successfully under other flow conditions is presented in Table 20. In this table, the first row of Case I

Table 20. Filter design to meet all wastewater flow conditions

Flow conditions	Q MGD	Backwash (10%) MGD	Filter capacity MGD	Nominal filter rate gpm/sq.ft.	Nominal[a] run length hr.	Nominal filter area sq.ft.	Actual filter area sq.ft.	Actual filter rate gpm/sq.ft.	Actual[a] run length hr.	Net water prod. gpd/sq.ft.[b]	Net water prod. MGD	Remarks
					I. C_0 = 40 mg/ℓ SS							
Max. 4-hr. peak (wet yr.)	18.1	1.81	19.91	6.25	8.0	2210	2940	4.70* 6.25**	10.5* 8.0**	6,250* 8,250**	18.4* >18.1 18.2**>18.1	Suff. Suff.
Max. daily (normal yr.)	11.3	1.13	12.43	3.90	12.5	2210	2940	2.93* 3.90**	15.7* 12.5**	3,930* 5,200**	11.5* >11.3 11.5**>11.3	Suff. Suff.
Mean annual	8.82	0.88	9.70	3.04	15.5	2210	2940	2.28* 3.04*	18.5* 15.5**	3,090* 4,080**	9.1* >8.82 9.02**>8.82	Suff. Suff.
Min. 4-hr. in low flow month	3.82	0.38	4.20	1.32	24.0	2210	2940	0.99* 1.32**	29.0* 24.0**	1,300* 1,750**	3.82* =3.82 3.86**>3.82	Suff. Suff.

* four filter cells in operation.
** one filter cell down for repairs.

[a]From Fig. 32.

[b]From Fig. 33.

Table 20. (continued)

Flow conditions	Q MGD	Backwash (10%) MGD	Filter capacity MGD	Nominal filter rate gpm/sq.ft.	Nominal[a] run length hr.	Nominal filter area sq.ft.	Actual filter area sq.ft.	Actual filter rate gpm/sq.ft.	Actual[a] run length hr.	Net water prod. gpd/sq.ft.[b]	Net water prod. MGD	Remarks
				II. C_o = 20 mg/ℓ SS								
Max. 4-hr. peak	18.1	1.81	19.91	6.25	14.0	2210	2940	4.70* 6.25**	21.5* 14.0**	6,500* 8,800**	19.2* >18.1 19.4**>18.1	Suff. Suff.
Max. daily (normal yr.)	11.3	1.13	12.43	3.90	23.0	2210	2940	2.93* 3.90**	29.5* 23.0**	4,050* 5,400**	11.8* >11.3 11.9**>11.3	Suff. Suff.
Mean annual	8.82	0.88	9.70	3.04	28.8	2210	2940	2.28* 3.04**	37.[illegible]* 28.8**	3,150* 4,200	9.26* >8.82 9.28**>8.82	Suff. Suff.
Min. 4-hr. in low flow month	3.82	0.38	4.20	1.32	60.0	2210	2940	0.99* 1.32**	70.0* 60.0**	1,350* 1,820**	3.96* >3.82 4.02**>3.82	Suff. Suff.

(C_o = 40 mg/ℓ) represents the 4-hr. peak flow on the maximum day transported from the third alternative of Table 19. The second, third and fourth rows of Table 20 represent the three flow conditions of maximum daily flow, mean annual flow and minimum 4-hr. flow in the low flow month. All the calculations presented in this table are arrived at in the same way as the calculations in Table 19.

The results for Case I (C_o = 40 mg/ℓ) show that this filter design is sufficient to filter the wastewater under expected flow conditions during the maximum 4-hr. peak, maximum daily, mean annual and minimum 4-hr. flow in the low flow month when all four filter cells are in operation and when one filter cell is out of service. A further check is required for conditions when there are low SS concentrations in the wastewater.

iii. Checking at low SS concentrations

Case II of Table 20 shows the calculations of net water production under various flow conditions when the influent SS concentration is assumed to be half of that assumed for design purposes (C_o = 20 mg/ℓ). The results shown in Case II of Table 20 indicates, as expected, that this design will operate successfully under all flow conditions with lower SS concentrations in the wastewater. Comparing the net water production of Case I (C_o = 40 mg/ℓ) and Case II (C_o = 20 mg/ℓ), it reveals that a filter design based on the worst

SS concentration will be sufficient when the SS concentration is lower than the designed level. It is interesting to note that when the SS are only half those used for design purposes, the run lengths under equivalent conditions are twice as long, but the net production obtained from the filter is increased only 3 to 4 per cent.

c. Characteristics of tertiary filter plant for Ames, Iowa

Number and type of filter: one CenTROL 4-cell gravity filter

Total filter area: 2940 sq. ft.

Media type, size, and depth:

Type	Size, mm	Depth, in.
anthracite	1.84 (10/12)	12 to 15
sand	0.55 (30/35)	12 to 15

Nominal filtration rate: 6.25 gpm/sq. ft. (actual filtration rate varies according to the flow condition, Table 20)

Expected run length: varies according to the flow conditions and levels of SS concentration (Table 20)

Backwashing: 3 min. air scouring with 3-5 scfm/sq. ft. followed by 5 min. water wash with rate of 20 gpm/sq. ft.

Availability: to meet all conditions of flow and SS concentrations

Expected capital cost: $300,000

6. <u>Design example 2: Using two or more 4-cell gravity CenTROL filters</u>

In practice, it is not usual to depend on a single treatment unit in a plant whose mean flow exceeds 1 MGD. In the previous example, the one CenTROL filter unit has 4 separate filter cells, but if some part of the unit malfunctions, it might be necessary to shut down all four filters. The use, therefore, of two CenTROL units would provide more versatility in plant operation and more safety against filter failure. In effect, the total filter area found to be required in design example 1 can be divided and placed in two or more CenTROL units. The cost of the filters would then be (from Fig. 35):

1 CenTROL unit	\$ 300,000
2 CenTROL units	\$ 360,000
3 CenTROL units	\$ 435,000

VII. SUMMARY AND CONCLUSIONS

The application of granular filters for tertiary wastewater treatment has resulted from the need for providing a higher treated effluent quality to meet more stringent stream water quality standards. Granular filters are currently finding application in tertiary treatment of biologically treated effluents where they are being used in plain filtration of the secondary effluent or as a last step in processes involving lime or alum treatment for phosphorous precipitation, settling and filtration. Granular filters are also an essential step in the physico-chemical treatment methods now being advocated as a potential competitive replacement for biological treatment of wastewater (94). The use of granular filters in these applications is being advocated even though many basic questions about the design and operation of such filters in wastewater treatment have not been answered.

This study was designed to answer some of the questions concerning the problems involved in wastewater filtration. The study was planned and conducted in six sequential phases with a specific purpose in each phase. They are: (1) phase A: effect of media size on filtration of wastewater through a single media sand filter; (2) phase B: effect of flow rate on filtration of wastewater through a single media sand filter; (3) phase C: comparison of operating characteristics of single media versus dual media filters using wastewater; (4) phase D: effect of size of anthracite and sand on filtration of wastewater through a dual media filter; (5) phase E: effect of media size on filtration of wastewater through a single media anthracite filter; (6) phase F: effect of flow rate on filtration of wastewater through dual media of selected size.

In a single media sand filter, surface clogging predominated and prevented the solids from penetrating into the lower layers of the filter bed and caused high headloss build-up. As a consequence, the removal capacity of the filter bed was not fully utilized. A more efficient filter design, an anthracite-sand dual media filter, was studied to investigate its superiority to the single media sand filter based on the filter effluent quality, headloss development and quantity of water produced.

A method was developed and demonstrated for selecting the size and depth of anthracite, used in the upper portion of the filter, and sand, used in the lower portion, taking into account the effect of intermixing. On the basis of the tests completed in phases A, B, C, D, and E, it has been found that for the filtration of the Ames final effluent wastewater a dual media filter should have 12 to 15 in. of uni-sized 1.84 mm anthracite on top of 12 to 15 in. of uni-sized 0.55 mm sand. With this filter characteristic, filtrate quality obtained was in the range of 5 mg/ℓ SS when the flow rate was as high as 6 gpm/sq. ft. and the influent suspended solids concentration was as high as 40.5 mg/ℓ. There appeared to be no significant effect on filtrate quality due to the high flow rate (Fig. 23), which confirms results described in the literature (86, 96). Headloss development was found to be related to the SS accumulation within the filter pores (Fig. 28) and was relatively unaffected by the filtration rate, which confirms the assumption made by Tchobanoglous and Eliassen (85).

Plain filtration of wastewater removes SS, BOD and other chemical pollutants. It is believed that the BOD removed by filtration is primarily the portion associated with the SS, that is, the portion of undissolved BOD. Results from a typical filter run showed that about 88 per cent SS removal and 45 per cent BOD_5 removal were obtained.

In analyzing the results from this pilot plant operation, special attention has been paid to investigation of the effect on filter performance of the variation in wastewater characteristics. The quality of the wastewater coming into the filter plant varies, primarily due to variations in treatment efficiency of the preceding units (primary, biological treatment and final clarifier). The variations of the characteristics of the wastewater change from time to time. This variation affects the filter removal mechanisms in two ways: (1) variation in the proportion of the total solids which are removed through surface removal versus depth removal mechanisms; and (2) variation in the proportion of the total solids whose removal is more affected by transport phenomena versus those whose removal is more affected by attachment phenomena. As a result, it is rather difficult, if not impossible, to predict the rate of exhaustion of SS removal and hydraulic capacities.

Therefore, prior to designing a tertiary filter plant, a pilot plant operation is essential if the plant is to approach optimum economic design. The pilot plant operation should be designed to:

(1) determine the proper size and depth of anthracite and sand based on that particular type of wastewater;

(2) determine the backwash rate required for the media size and depth combinations as determined in (1) which will expand the media bed to the desired degree;

(3) develop the curves relating the net water production to the filtration rate at various run length conditions (Fig. 33); and

(4) develop the curves relating the run length to the filtration rate at various influent SS concentrations (Fig. 32).

Based on the information generated from this pilot plant study, a case study for designing a tertiary filter plant for the Ames pollution control plant was made. Several conclusions were drawn from this case study. They are:

(1) There exists an upper limit of net water production for each filtration rate. As shown in Fig. 33, the upper limits of net water production for filtration rates of 2, 4 and 6 gpm/sq. ft. are 2,880, 5,760 and 8,640 gpd/sq. ft., respectively, which is the net water production from a theoretical infinite run length, i.e. no backwash water required. Information from Fig. 33 reveals that a shorter run length at a higher filtration rate can produce the same amount of filtrate that a long run length at a low filtration rate can. For example, a run length of 20 hr. at 2 gpm/sq. ft. produces 2,700 gpd/sq. ft. filtrate. This same amount can be produced with a 1 hr. run length at 4.5 gpm/sq. ft.

(2) The nominal filtration rate should be determined based on the maximum 4-hr. peak flow and worst SS concentration in the wastewater. The net water production to be provided by the filters should be equal to or greater than the design plant flow

(maximum 4-hr. peak flow) when either all four filter cells are in operation or one of the filter cells is down for repairs or backwashing. A trial and error method can be used, as shown in Table 19, to determine the nominal plant filtration design rate. The nominal filtration rate for a particular filter plant depends on the peak flow condition, SS characteristics and concentrations in the wastewater to be filtered and the physical characteristics and arrangement of the filter cells (number of cells in a filter unit).

(3) The filter which has been designed based on the maximum 4-hr. peak flow condition should be checked to see whether it is sufficient at other flow conditions, such as the maximum daily, mean annual and minimum 4-hr. in the low flow month (Table 20).

(4) The filter design should also be checked for its capacity of water production under SS concentrations other than the designed level (Table 20).

In conclusion, this study has demonstrated that a granular filter can be designed to treat secondary final effluent successfully, provided the necessary information is generated from a pilot plant operation and the rational design method as demonstrated in this thesis is followed.

VIII. RECOMMENDATIONS

Due to the fact that there are complicated interrelationships among the many factors affecting filter performance, besides the variations in the characteristics of the wastewater delivered to the tertiary filter plant, a pilot plant operation should be conducted prior to actual design of a filter for tertiary treatment. A pilot plant study for a particular wastewater should be designed to:

1. determine the proper anthracite size and depth in dual media filters (phases C and E).
2. determine the proper sand size and depth in dual media filters, taking into consideration the effect of intermixing (phase D).
3. determine the backwash rate required for the media size and depth combinations as determined in 1 and 2;
4. develop the curves relating the net water production to the flow rate at various run length conditions (Fig. 33); and
5. develop the curves relating the run length to the flow rate at various influent SS concentrations (Fig. 32).

Based on the information generated from the pilot plant operation for the particular wastewater to be filtered and the rational design procedure demonstrated in this thesis, a granular filter can be designed and operated successfully.

The following are studies which need to be made to extend and amplify the results of this dissertation:

1. A study should be conducted of the effect of various degrees of intermixing, which is determined by the media size ratio between layers, on the filter performance. It is expected that the degree of intermixing will affect the efficiency of the solids removal and the rate of headloss build-up. The determination of the optimum degree of intermixing in a dual media filter is an essential step in a dual media filter design.
2. Study is required to investigate the feasibility of and to develop a wastewater filtrability index for correlating wastewater characteristics with granular filter design criteria for tertiary wastewater treatment.
3. Throughout the course of the pilot plant operation, no sign of solids breakthrough occurred even when the headloss was as high as 24 ft. of water.
   A further study should be conducted to investigate whether solids breakthrough will occur at much higher headloss conditions. Thus, pressure filters may be applicable for wastewater filtration, since pressure filters can be designed for much higher

headloss conditions without increasing the structural cost as in the case of a gravity filter. The comparison of filtration costs between pressure and gravity filters should be made such that a more economical type of filter for tertiary wastewater treatment can be determined.

4. The most immediate study required is to investigate the effect of various influent solids concentrations on the rate of exhaustion of SS removal and filter hydraulic capacities. Then, a practical mathematical model to predict wastewater filter performance can be obtained. With such information, more sophisticated techniques of optimization such as linear programming and/or dynamic programming can be applied to optimize the system of wastewater filtration.

5. The studies described in this thesis were made without chemical treatment of a secondary plant effluent. Similar studies should be made using chemically treated effluents.

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X. ACKNOWLEDGMENTS

The writer would like to express his appreciation to his major professor, Dr. E. R. Baumann, for the guidance provided throughout his entire program of graduate study and research.

The writer would like to thank Dr. Harris F. Seidel, Director of Water and Pollution Control, Ames, Iowa, for permission to conduct this study at the Ames water pollution control plant, and Mr. Ralph Briley and Mr. Milo Sampson, Chief Operators at the plant, for their help and cooperation.

Thanks also go to Mr. Dave Millard, Instructor in Engineering Extension, for his help in setting up the electric system for the pilot plant.

Special thanks are given to Mr. Forest R. Kenny of the General Filter Company for providing detailed cost information for Advanced Waste Treatment CenTROL Filters.

This research was supported by the Iowa Engineering Research Institute in part through funds made available by the City of Ames, Iowa, and Research Fellowship 5-FI-WP-26, 384-03 of the Water Quality Office, Environmental Protection Agency.

Lastly, the writer would like to thank his wife, who has helped in many ways during his academic program.

XI. APPENDIX A. SUMMARY OF GRANULAR FILTERS FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft²)	Max. terminal head (ft water)	Run length (hr)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Remarks
Dr. Mohr in Germany Rapid Gravity filter (from Zack[97], 1937)	Primary settled effluents	Sand 1~2		1.47	4.1	8								
	"	Sand 2~3		0.82	4.1	25								
Atlanta, reported by A. Potter (from Zack[97], 1937).	Chem. treated primary effluent	Anthracite E.S. 0.45		2	5	12	0~75		79				twice as many backwashings required for influent of 57 ppm as for 19 ppm SS.	
		"		3.5	5	3	9~75		79					
Wuppertal pilot filter (German) (from Streander[83], 1940)	Primary settled influent	Sand 3~4	28			40~50								low efficiency. Reduction 12% of oxygen consumption.
	"	Sand 2~3	12	0.75~1.5	3.9	20~25								Reduction 12 to 21% of oxygen consumption.
	"	Sand 1~2	28	0.75~1.5	3.9	16~8			80~90				Prior to the advent of high velocity wash mechanical rakes and low velocity backwash or air agitation followed by water washing	Reduction in O_2 Consumed: 20% for 1.5 gpm/ft² 25-30% for 1.1 gpm/ft² 35% for 0.75 gpm/ft² Retained solids well distributed throughout the bed
Wuppertal full scale plant 26' 3" x 123' (from Streander[83], 1940)	Primary settled effluent	Sand 1~2	28	0.83	4.25				40				Air wash, supplied by centrifugal blower - 2.4 ft³/min/ft². The rate of backwash in a deep sand filter is controlled by the size used, 6 ~ 8 in/min.	Low removed by SS was due to the inability to thoroughly remove the retained solids from the sand bed. Backward rate used was too low.

Source Information	Wastewater Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Remarks
South River plant, N. J. (from Streander[83], 1940)	Primary settled effluent	Sand 1.0 U.C. 1.5	6	0.64		1.9	132	70	47			28	37% treated water used	Total washing time/day = 190 min.
Sayerville, N. J. (from Streander[83], 1940)	Primary settled effluent	Sand 1.0 U.C. 1.5	6	0.6		2.7	154	76	51			13	47% treated water used	Total washing time/day = 135 min.
Sayerville, N. J. (from Streander[83], 1940)	Chem. treated primary settled effluent	Sand 1.0 U.C. 1.5	6	0.6		3	40	20	50			9	41% treated with water	Total washing time/day = 120 min.
Laughlin filter, Atlantic City (from Streander[83], 1940)	Primary settled effluent	Sand, E.S. 0.4 ~ 0.5 U.C. 1.8 ~ 2.0	6	2			88	39	56	187	137	26		Reduction in BOD is not proportional to reduction of SS
Aucor Sewage Works, pilot plant, South Africa (from Vosloo[93], 1947).	Secondary settled effluent	Sand 0.5 ~ 1.7	29	2	10	6 ~ 18	24.3	0.7	97				5% of total flow used for backwash	
Pilot plant at Luton, England 1949-1950 (from Pettet, Collett, Waddington[70], 1952)	"	Sand 0.85 ~ 1.7	24	2	7	24	24	3.4	86	26	10	61.5	2 ~ 3% of treated water, 13.3 gpm/ft^2 rate	No appreciable difference in head loss for sand filters with 2 ft and 3 ft 6 in depth, respectively. Also little difference in effluents from two filters at rates below 2.92 gpm/ft^2, but at higher rates, the effluent from filters of 3'-6" of sand was slightly superior.
	"	Anthracite 1 ~ 2	24	2	7	24	25	3.3	87	26	9	65.5	2 ~ 3% of treated water, 10 gpm/ ft^2 rate	
	"	Sand 0.85 ~ 2	24	2.34 ~ 3.42			19.3	0.9	95	23.8	7.2	70		
		"	42	2.34 ~ 3.42			19.3	0.7	96	23.8	6.3	74		

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft²)	Max. terminal head (ft water)	Run length (hr)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Remarks
Luton Plant, England May 1951 ~ April 1952 (from Evans, Roberts[19], (1952)	Secondary final effluent	Sand 0.85 ~ 1.67	8	3.33	8~9	24	15.4	4.2	73	13.0	6.7	49	By air-scour, 2.5% of treated water	1. During storm flow and high loading, backwash every 12 hrs. 2. Lower DO in filter effluents. 3. Oxidation of ammonia range from 35 to 68%. 4. Cost per lb of SS removed = 4.25 d Cost per lb of BOD removed = 7 d. 5. Visible color reduce 6.5%; thiocyanate, 50% phenals, 8%.
Full scale sand filter at Luton, England (From Evans[17], 1957).				2.5		24	13.4	3.3	76	9.2	4.0	56.5	Use 3% of treated water	
Pretoria Sewage-treatment works, Johannesburg, South Africa (from Nicolle[64], 1955; Nicolle[65], 1957)	Secondary final effluent	Sand 0.84 ~ 0.59		3	6.5	8~10	22	4.6	79				Air-water rate: 25% gpm/ft²	
		1.40 ~ 0.65		3	6.5	8~10		8					9.5% of treated water used	
Los Angeles Hyperion treat-ment plant (preliminary tests)(from Laverty et al.[45], 1961).	Activated sludge effluent	Sand E.S. 0.95 U.C. 1.6	11	2			26.1	12.7	46	9.8	4.8	51	automatic backwash	

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Remarks
Lab study at Peoria Sanitary District Treatment Works (from Fall & Kraus[21] 1964).	high rate activated sludge effluent			1.12			16.0	8	50	28.3	17	40		Filter clogged easily and run length was only a few hours.
Research at Iowa State University. Pilot plant was located at Ames Sewage treatment plant. (Merry[57], 1965)	Trickling filter final effluent	Sand E.S. 0.55 U.C. 2.36	24	2	6.25	10	19.8	5.7	71.2	56	24.2	56.8	No air wash 20-25 gpm/ft^2 for 10 min.	Clogged filters might have not been cleaned properly
				4		6.25	18.5	6.0	67.6	52.6	23.4	55.5		
				6		4.67	17.7	6.3	64.4	50.4	24.4	51.6		
Philomath Municipal sewage treatment plant, Oregon (from Culp, Hansen[12], 1967)	Extended aeration effluent	mixed-media	30	5			59	4	93	26	2.5	90	Automatic control, No air wash	Remove phosphate by adding chemical
Gravity sand filter and pilot up-flow filter at Luton, England 1966 (from Naylor, Evans, and Dunscombe[63], 1967)	final settled (down-flow full scale)	Sand 0.85~1.7	36	4.0			12.8	7.5	41	5.3	3.1	42		up-flow filter produced a better effluent.
	(up-flow pilot plant)	Sand 0.85~1.7	60	4.2			12.8	5.7	56	5.3	2.6	51		
Cambridge, England, 1967 (by Holden Discussion[24], 1967)	final settled effluent (down-flow)	Sand 0.85~1.7	24	2.5	3	6	20-40		77					1. Tertiary treatment could only be successful when the influent was well oxidized 2. Effluent poorer at higher rate
	(up-flow)	Sand 0.85~1.7	24	2.5		24	20-40							

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5(mg/l) Infl.	BOD_5(mg/l) Effl.	BOD_5(mg/l) % Removed	Backwash Information	Remarks
West Hertfordshire Authority, England, 1968 (from Wood, Smith, Murry[95] 1968)	final effluent of trickling filter (up-flow)	Sand 1~2	60	1.8	5		44.3	1.9	95.7	57.7	3.9	93.1	Air scour then backwash with untreated effluent	1. Sharp fall in efficiency for flow greater than 4 gpm/ft^2 2. Little or no nitrification or denitrification took place in the filter
		Sand 1~2	60	3.33	5		37.3	3.7	90.1	53	4.6	91.3		
		Sand 1~2	60	4.16	5		55.5	7.1	87.1	42	5.6	86.6		
		Sand 1~2	60	5.0	5		37.3	9.9	73.5	34.6	14.7	57.5		
Letchworth Plant, England 1966 ~ 1967 (from Truesdale and Birkbeck[91] 1968)	final effluent of activated sludge (up-flow)	Sand 1~2	60	4.4			17.2	6.9	60	19.2	9.0	53	Air scour then washed by upward flow to about 11.7 gpm/ft^2. 6% treated water used	Effluent from filter was inferior at higher rates. (>6.6 gpm/ft^2)
Lebanon pilot plant, Ohio (from Berg & Brunner[2], 1969	Activated sludge final effluent with polyelectrolyte (2.5 mg/ℓ) and/or alum (12 mg/ℓ)	Single coal E.S. 0.75 U.C. 1.5	20	5	28.3	3~6	5~25						1. rate of 15~20 gpm/ft^2 2. backwash water used 6.6% (for 2.5 hr run) 11% (for 1.5 hr run) of treated water	1. Effluent quality less than 0.5 JTU 2. Total cost 6.7c/1000 gal. 50% of cost for electric power and chemicals.
				10		1.5~3								
		dual-media 14" coal E.S. 0.75, U.C. 1.5 6" sand E.S. 0.75, U.C. 1.6		5	28.3	2.5~4								
				10										
Hanover treatment plant Metro Chicago, Harding design filter (from Sosewitz & Bacon[78] 1968, from Lynam, Ettelt, & McAloon[55], 1969; from Lynam, Bacon[54], 1970)	final effluent of activated sludge	Sand E.S. 0.51 U.C. 1.62	12			4~5			77			89	traveling backwash	1. Efficiency of tertiery treatment depend on the efficiency of secondary treatment. 2. Additional solids removal provided by coagulation with alum plus polymer was not sufficient to warrant inclusion with filtration into tertiery treatment design.

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Remark
Pilot scale sand filter at Derby, England (from Oakley & Cripps[66], 1969; Joslin & Green[39], 1970)	Trickling filter final effluent	comm. sand 1.2 ~ 2.4	24	2.5	12	14-17			59-60			23-38		
				3.33	12	20			57-68			50-64		
				5.0	12	17-18			47-58			32-37		
		comm. sand 1.2 ~ 2.4	36	3.33	12	19			65			67		
				5.0	12	16-17			60			43-46		
		L.B. sand 1.2 ~ 1.68	24	2.5	12	10-12			59-60			54		
				3.33	12	11-15			52-64			53-65		
	Trickling filter final effluent	Three-layer filter Anthracite 8"-2.4 ~ 1.5		2.5	12	13			64			69		
		Sand 8"-1.5 ~ 1.19		3.33	12	13-17			57-70			59-71		
		Garnet 8"-0.85 ~ 0.7		5.0	12	12-14			53-65			34-46		
		Sand 4'-1 ~ 2 Gravel 2'-2" (up-flow)		3.33	12	25-26			57-74			60-67		
				5.0	12	12-17			47-58			48-50		
High Rate filtration of Mill Scale waste (from Donovan[13],	Mill scale waste	Anthracite 5.1	84	8	15	46	150	11	92.5				Air scour for 5 min. at 8 cfm/sf then flush with water at 30 gpm/sf for 10 min.	The cationic polyelectrolyte produced the best results. The cost of feeding polyelectrolyte was less than $1.00 per million gallons water filtered.
				16	15	38	150	36	76					
				23	15	33	150	83	45					
				30	15	23	150	85	43					
		Sand 2-3	84	8	15	20	150	2	99					
				16	15	9.5	150	4	98					
				23	15	20.5	150	30	80					
				30	15	3.5	150	7	95					
		Anthracite 4'-5.1 Sand 3'-2.3		8	15	28	150	3	98					
				16	15	20.5	150	3	98					
				23	15	12	150	10	93					
				30	15	8.5	150	9	94					

XII. APPENDIX B. DETERMINATION OF THEORETICAL BACKWASHING TIME

Backwash time is defined as the total period of time when a filter is down for backwashing, including the time for (1) draining down the water above the filter media to a level just below wash trough; (2) air scouring; (3) water washing; and (4) leeway. It has been common practice to estimate the backwash time in the filtration cycle to be a half hour. However, the backwash time depends on the flow rate used as illustrated in the following example:

In this example, the terminal headloss is assumed to be 10 ft. water, which is the water column above the surface of the filter media, and the wash trough level is assumed to be 4 ft. above the surface of the filter media, as shown in Fig. 39. When the headloss reaches the predetermined 10 ft. water level, the filtration cycle is terminated and the backwashing cycle is started. Before initiating the air scouring, the residual water above the filter media is filtered down to the level of the wash trough in order not to waste this water. The time period required depends on the filtration rate used. If a uniform decreasing of the draining rate from the terminal headloss water level to the surface of the filter media is assumed, an average draining rate can be estimated and the time required for filtering down to the level of the wash trough is obtained as follows:

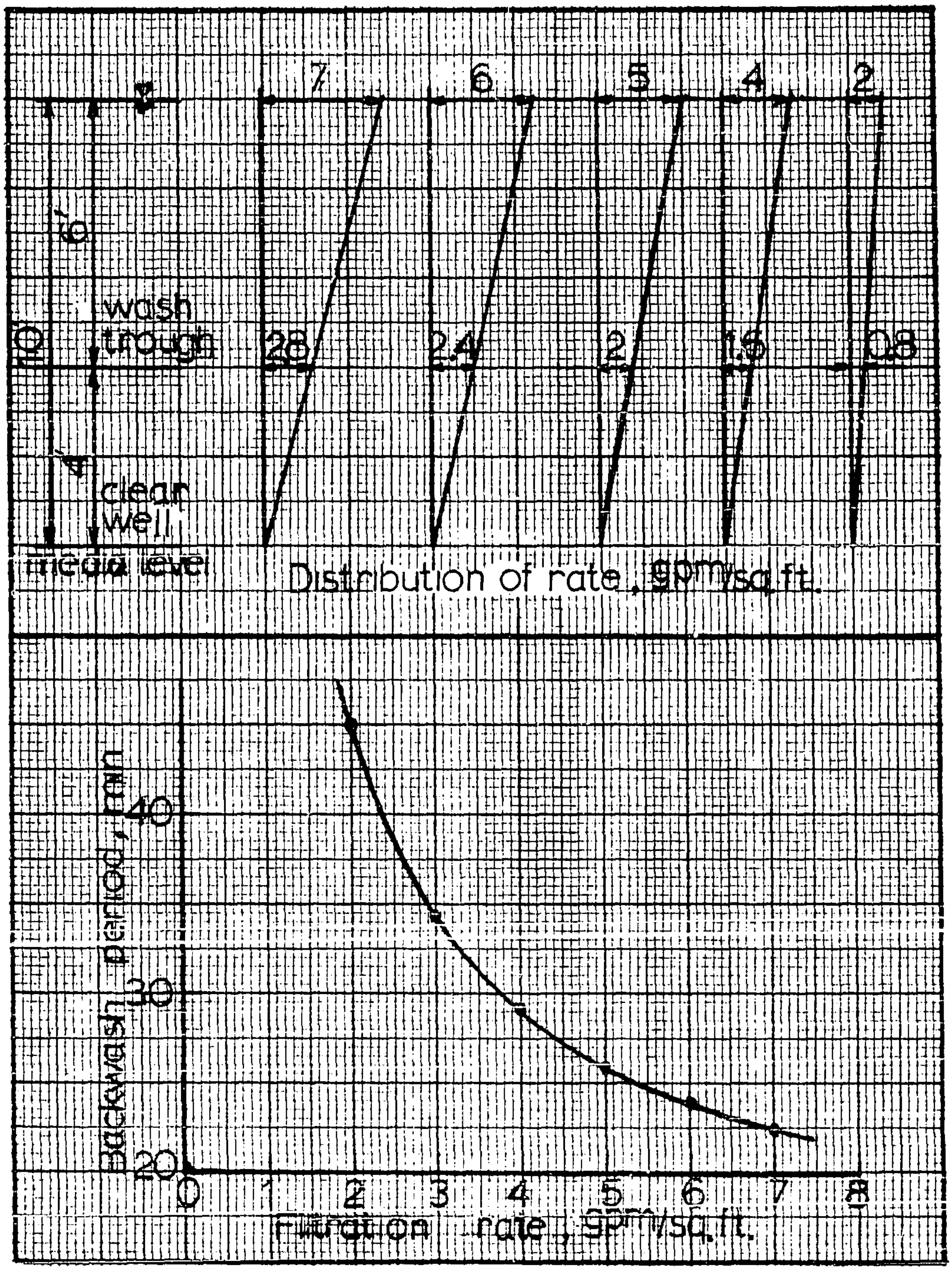

Fig. 39. Theoretical backwashing period vs. filtration rate

For a filtration rate of 7 gpm/sq. ft.:

Average draining rate =

$$\frac{7(\text{gpm/sq.ft.}) + 2.8(\text{gpm/sq.ft.})}{2} = 4.9 \text{ gpm/sq.ft.}$$

Time required for draining =

$$\frac{6(\text{cu.ft.}) \times 7.48(\text{gal/cu.ft.})}{4.9(\text{gpm/sq.ft.})} = 9.15 \text{ min.}$$

Therefore, total down time will be

drain time	9.15 min.
air scouring	3.00 min.
water washing	5.00 min.
leeway	5.00 min.
	22.15 min.

A similar analysis can be made for cases of filtration rates other than 7 gpm/sq. ft. The results are shown in the lower portion of Fig. 39. The results from Fig. 13 indicate that the period required for filter backwashing increases exponentially as the filtration rate decreases. The typical periods of down time are:

Filtration rate gpm/sq. ft.	Average drain rate gpm/sq. ft.	Down time min.
2	1.4	45.0
3	2.1	34.4
4	2.8	29.0
5	3.5	25.8
6	4.2	23.7
7	4.9	22.15

The relationship between net water production and filtration rate at various run lengths using theoretical backwashing time is shown in Fig. 40.

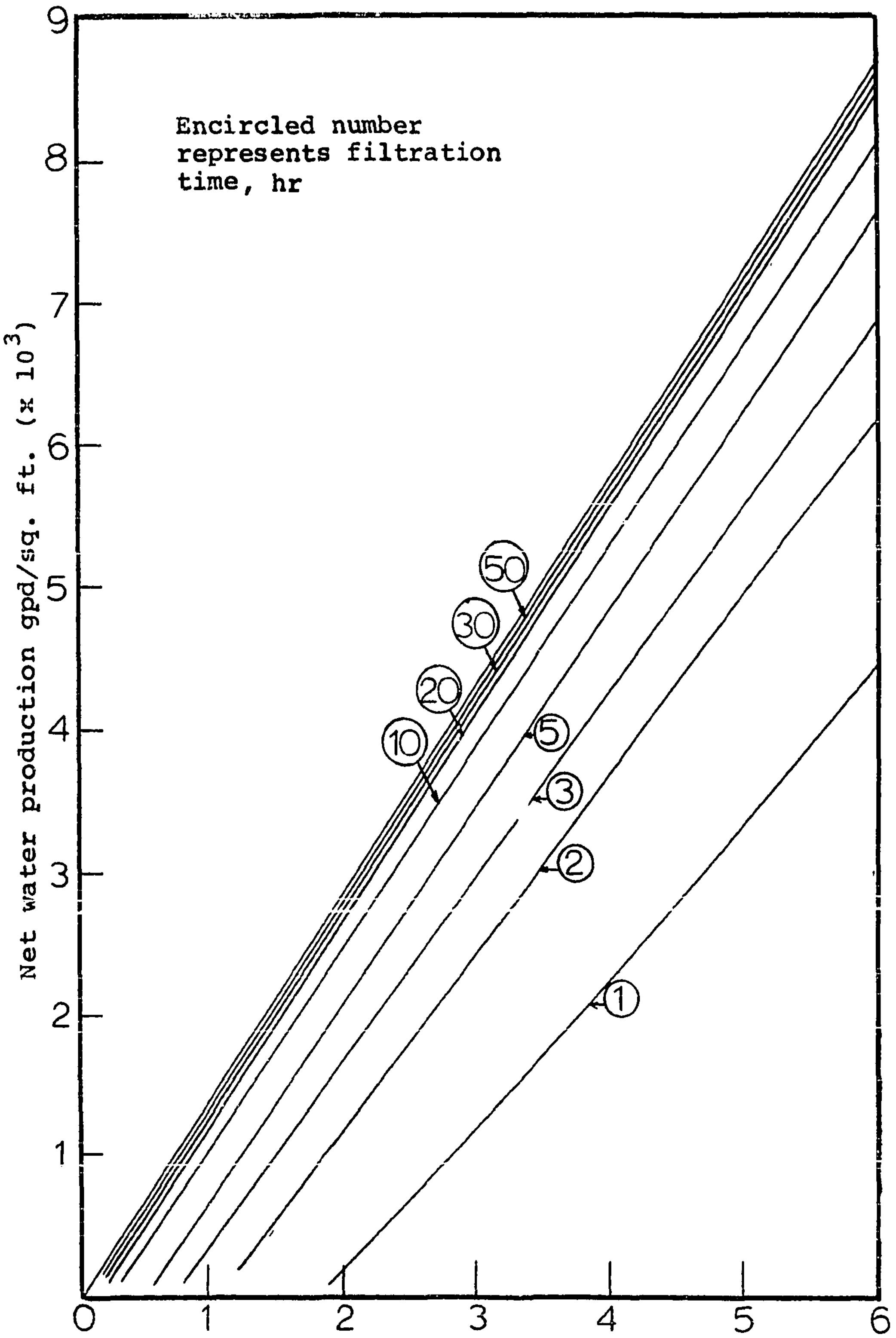

Fig. 40. Net water production vs. filtration rate at various run lengths (using theoretical backwashing time)

XIII. APPENDIX C. FILTER BED EXPANSION VS. BACKWASH RATE

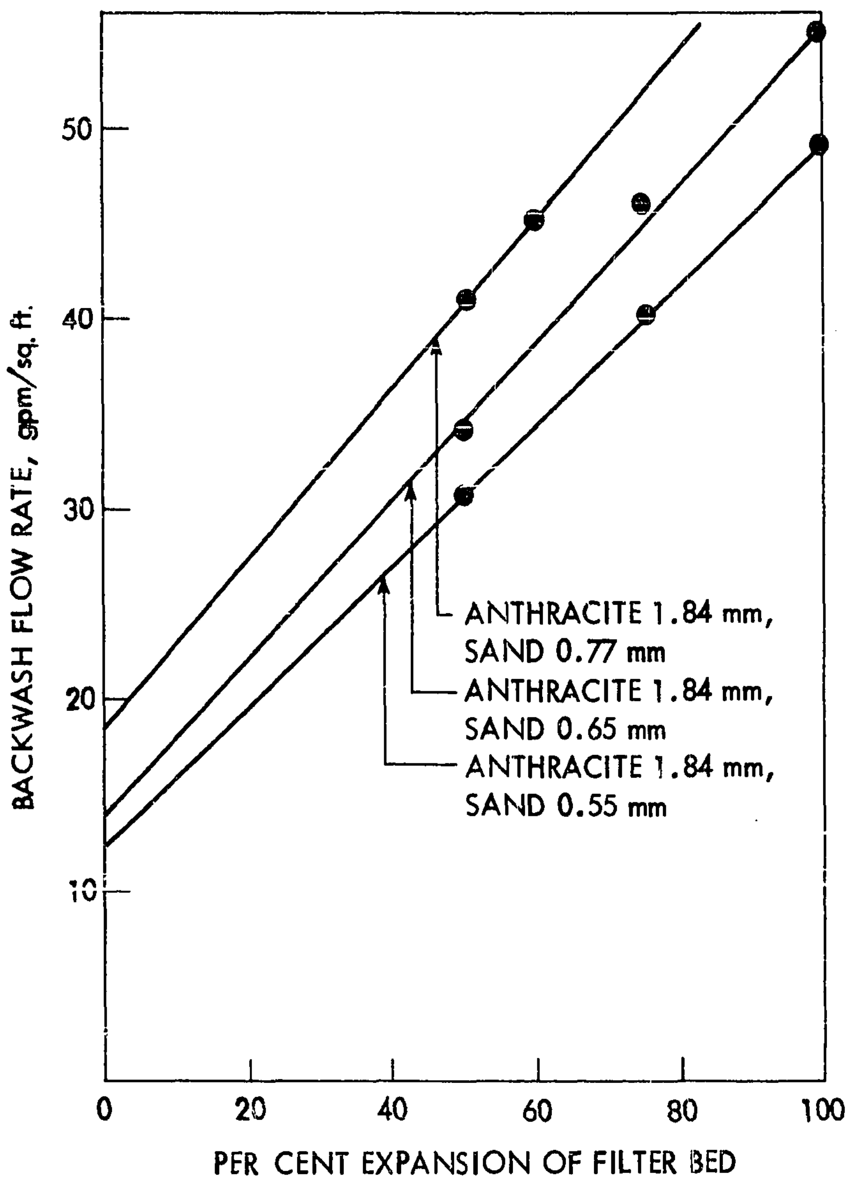

Fig. 41. Media bed per cent expansion vs. backwash flow rate at various media size combinations (water temperature 50-60°F)