ROBERT A. TAFT WATER RESEARCH CENTER

REPORT NO. TWRC-11

APPRAISAL OF GRANULAR CARBON CONTACTING

PHASE I

EVALUATION OF THE LITERATURE ON THE USE OF GRANULAR CARBON FOR TERTIARY WASTE WATER TREATMENT

PHASE II

ECONOMIC EFFECT OF DESIGN VARIABLES

ADVANCED WASTE TREATMENT RESEARCH LABORATORY - XI

U.S. DEPARTMENT OF THE INTERIOR

FEDERAL WATER POLLUTION CONTROL ADMINISTRATION

OHIO BASIN REGION

Cincinnati, Ohio

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Produced under the auspices of
The M. W. Kellogg Company, A Division of Pullman Incorporated

for

THE ADVANCED WASTE TREATMENT RESEARCH LABORATORY
Robert A. Taft Water Research Center

This report, the first two phases of a three phase program, is submitted in partial fulfillment of Contract No. 14-12-105 between the Federal Water Pollution Control Administration and the Swindell-Dressler Company, A Division of Pullman Incorporated.

U. S. Department of the Interior
Federal Water Pollution Control Administration
Cincinnati, Ohio

MAY 1969

FOREWORD

In its assigned function as the Nation's principal natural resource agency, the United States Department of the Interior bears a special obligation to ensure that our expendable resources are conserved, that renewable resources are managed to produce optimum yields, and that all resources contribute their full measure to the progress, prosperity, and security of America -- now and in the future.

This series of reports has been established to present the results of intramural and contract research studies carried out under the guidance of the technical staff of the FWPCA Robert A. Taft Water Research Center for the purpose of developing new or improved wastewater treatment methods. Included is work conducted under cooperative and contractual agreements with Federal, state, and local agencies, research institutions, and industrial organizations. The reports are published essentially as submitted by the investigators. The ideas and conclusions presented are, therefore, those of the investigators and not necessarily those of the FWPCA.

Reports in this series will be distributed as supplies permit. Requests should be sent to the Office of Information, Ohio Basin Region, Federal Water Pollution Control Administration, 4676 Columbia Parkway, Cincinnati, Ohio 45226.

The total resources of Pullman Incorporated were made available for the execution of this program. The first two phases, included herein, were primarily the responsibility of the M. W. Kellogg Division with support provided by the Swindell-Dressler Division. The principal engineering staff members involved included A. E. Cover, L. J. Pieroni and E. V. Rymer. Coordination and participation were provided by J. F. Skelly, Project Director, and C. D. Wood, Project Manager. The third phase, "Engineering Design and Cost Estimate of a Granular Carbon Tertiary Waste Water Treatment Plant," is published separately as TWRC-12.

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ABSTRACT

A literature review of the data on tertiary waste water treatment has been made with a view towards generating sufficient basic data for process designs of various schemes for activated carbon adsorption of COD (Chemical Oxygen Demand) from waste water. Particular attention was given to the allowable capacity (loading) of carbon with organic waste matter and the effect of liquid linear velocity, carbon particle size, and number of regneration cycles on adsorption capacity and rate.

Findings of the work on tertiary waste water treatment using granular activated carbon are summarized. The economic importance of several design variables are determined and additional experimental data requirements identified. The variables, which were studied for their effect on the economics, were: shop fabrication and field erection of vessels, surge designs, plant size, idle carbon inventory, velocity, contact time, particle size, regeneration loss, carbon capacity, downflow, upflow, gravity-flow and moving-bed contactors, adsorbent cost, number of contacting stages, in-place regeneration, and certain combinations of the above variables.

Recommendations are made for further evaluation and experimental work.

The principal source of the data utilized in this project was the granular activated carbon treatment pilot plant operated at Pomona, California by the Los Angeles County Sanitation District for the Federal Water Pollution Control Administration. Additional information was collected from carbon treatment facilities located at Lake Tahoe, California, Nitro, West Virginia, Washington, New Jersey and Lebanon, Ohio. Other data was obtained from published and unpublished sources as listed under References.

PHASE I. EVALUATION OF THE LITERATURE ON THE USE OF GRANULAR CARBON FOR TERTIARY WASTE WATER TREATMENT

by

A. E. Cover and L. J. Pieroni, The M. W. Kellogg Company

SUMMARY

A literature review of the data on tertiary waste water treatment has been made with a view towards generating sufficient basic data for process designs of various schemes for activated carbon adsorption of COD (Chemical Oxygen Demand) from waste water. In this review, particular attention was given to the following critical process parameters:

1. Allowable capacity (or loading) of carbon with organic waste matter (reported as pounds of COD removed per pound of carbon) to purify water to an acceptable level. An allowable loading of 0.87 lb COD/lb carbon was chosen, based on Pomona pilot plant data for fixed bed, downflow systems assuming feed water characteristics comparable to those found at Pomona.

2. Effect of liquid linear velocity on adsorption capacity (loading and rate.) The literature data on the effect of linear velocity on adsorption rate are conflicting. Some of these data indicate an increase in adsorption rate with velocity. If this is true, reductions in investment could be realized by running at high velocities. More experimental data are needed in order to completely resolve this question.

3. Effect of carbon particle size on adsorption capacity and rate. It was found that at relatively short contact times (10 minutes) there will be a reduction in allowable capacity (loading) of 20% to 35% in going from 12 x 40 mesh to 8 x 30 mesh carbon based on Lake Tahoe data. Although there are incomplete data to support this reduction there should be less of an effect of particle size on capacity at the longer contact times required in commercial plants (up to 50 minutes).

4. Effect of number of regeneration cycles on adsorption capacity and rate. Carbon's ability to remove COD to the required concentration is not affected by regeneration but the carbon's capacity is decreased to 65% of its original capacity after seven regeneration cycles based on Pomona pilot plant data in a single contactor. However, the capacity still appears to be decreasing after seven regeneration cycles.

It is felt that the data are not fundamental enough in nature to permit their use as basic design data for system configurations other than the type in which they were taken.

However, based on these data, preparation of preliminary process designs and evaluations for the various types of proposed contacting systems should begin. Such studies will hopefully determine which scheme has the maximum technical and economic potential, and will indicate which experimental data - if any - are required to bring the selected scheme to commercial reality.

In view of the deficiencies in the present data, there are several gaps which must be filled before a precise economic optimization and comparison can be made of the several contacting systems. Recommendations are made for the following experimental work:

1. A side-by-side experimental comparison of upflow and downflow columns should be made to determine if there is any inherent inefficiency in either of these methods of contacting.

2. The effect of high and low velocities (up to 10 GPM per square foot and below 4 GPM per square foot) on rates of COD adsorption must be determined at long contact times. The effect of low velocities is essential if gravity flow systems are to be evaluated.

3. The effect of carbon particle size on adsorption capacity and rate before and after regeneration should be determined at contact times of commercial interest (up to 50 minutes).

4. The effect of the number of regeneration cycles on adsorption capacity and rate should be investigated to determine if the decreasing trend will eventually level off. The effect of velocity after regeneration should also be studied.

5. The effect of particle size on regeneration loss should be determined under closely controlled conditions.

INTRODUCTION

The purpose of this report is to summarize findings on the first portion of the work on tertiary waste water treatment with activated carbon. Specifically, the work reported herein represents the results of a thorough literature review of the data on tertiary waste water treatment with a view to generating sufficient basic data for process design of various schemes for carbon adsorption. In the literature review, particular attention was given to the following critical process parameters, the effects of which must be defined prior to any design work.

1. Allowable loading of carbon with organic matter to purify water to an acceptable level.
2. Effect of liquid linear velocity on adsorption capacity and rate.
3. Effect of carbon particle size on adsorption capacity and rate.
4. Effect of regeneration cycles on adsorption capacity and rate.

As a minimum result, the data extracted from the various sources should provide sufficient information to make empirical designs of plants of the type that have already been demonstrated (e.g., at Pomona or Lake Tahoe). Ideally, this survey would furnish basic design data which could be applied to any type of contacting scheme such as moving bed, upflow or downflow systems or fluidized bed systems, whether or not such a system has been piloted.

ALLOWABLE CAPACITY OF THE CARBON

A capacity, or loading, of 0.87 pound COD (total) per pound carbon or 0.58 pound dissolved COD per pound carbon will be used in design for fixed-bed downflow contractors operating on 60 mg/l COD (total) influent with feed water characteristics the same as at Pomona, based on results from the Pomona pilot plant (Figure 1) under the following conditions:

1. 7 GPM/ft^2 superficial velocity
2. 16 x 40 mesh, type CAL Pittsburgh Activated Carbon
3. Virgin carbon (unregenerated)

The effect of these three variables on the capacity will be discussed in subsequent sections.

This capacity (0.87 lb COD/lb carbon) will be used in design even though the design influent (60 mg/l) is higher than the average at Pomona (47 mg/l). The higher influent concentration should increase the loading if the displacement from exhaustion is great; however, this will not be considered here since no quantitative estimate of its effect can be made. Data needed for this type of correction would be operation of a column for an extended period on water with a higher influent concentration.

Adsorption isotherms are of no value here because the apparent capacity of the carbon in an operating column is 50% to 100% greater than the isotherm capacity.[1] This apparent increase in loading is probably due to biological activity. This result could be expected since the BOD is concentrated on the carbon and the rate of biological action should increase accordingly. Use of this uncorrected capacity implies that there will be a factor of safety but of undetermined magnitude.

The other runs shown in the Pomona report (Figure 1) with higher loadings (1.06 and 1.22 lb COD/lb carbon) were not producing a high quality effluent (see January-April 1966, Figure 2) and were discarded. However, these higher loadings do indicate that 0.87 lb COD/lb carbon is far from exhaustion.

The top line of Figure 1 is the contactor identification number. The designation, IV 0, A, B, C, D indicates that the contactor (IV) occupies the fourth position on the concrete slab, (0) that the carbon contained in the contactor has never been regenerated, and (A, B, C, D) that the contactor has occupied the first, second, third, and last position in a total column of four contactors.

Figure 2 graphically portrays the concentration of COD removed by each contactor within the four stage contactor column during the period June, 1965, - August, 1966, utilizing the same contactor identification number system as described for Figure 1.

	I OA		II OA,B		III OA,B,C		IV OA,B,C,D	
VOLUME TREATED (MG)	24		59		81		100	
DAYS ON STREAM	86		211		288		365	
DAYS IN POS'N. "A"	86		125		77		77	
CARBON DOSAGE (lbs/MG)	280		220		250		270	
WT. OF ORGANICS REMOVED	T	D	T	D	T	D	T	D
COD lbs/100 lbs of carbon	73	46	122	66	106	57	87	58
TOC " " " " "	28	–	32	16	27	18	24	17
ABS " " " " "		5.3		5.1		4.9		2.8

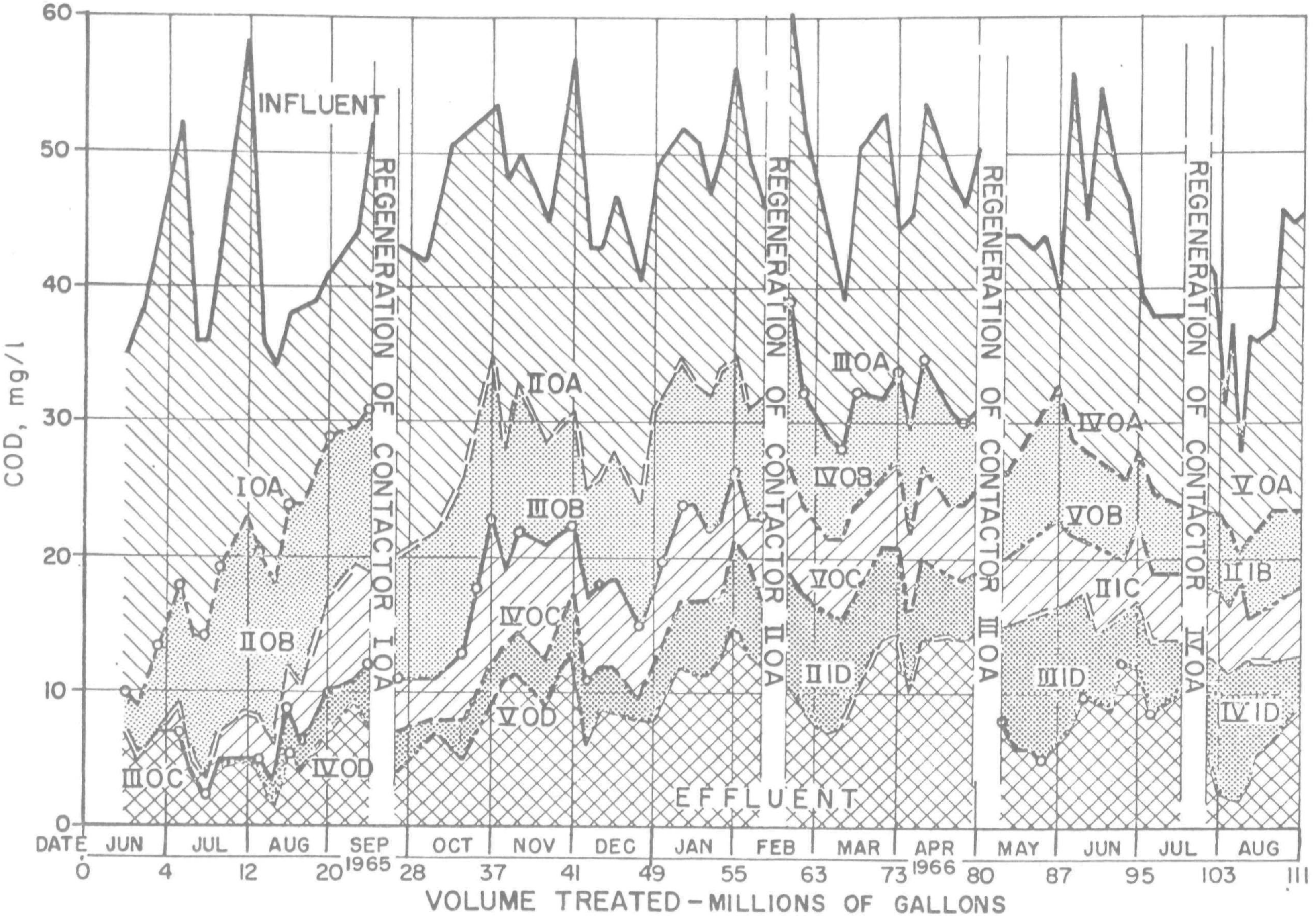
T = TOTAL
D = DISSOLVED

REFERENCE 2

FIGURE I

MAIN CARBON COLUMN PERFORMANCE

REPRINTED BY PERMISSION, Parkhurst, J.D., *et al.*, *JWPCF*, 39, Part 2, R77 (1967).
and County Sanitation Districts of Los Angeles County



REFERENCE 3

FIGURE 2

COD REMOVAL PATTERNS

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and County Sanitation Districts of Los Angeles County

The run chosen represents the only one where the contactor has been in all four positions in the total column before regeneration. This contactor was onstream for one year and was subjected to seasonal variations in influent concentration (Figure 2) which tend to average out this effect. This capacity also agrees with the recommendation in the Pomona report of 55-60 lb dissolved COD/100 lb carbon since dissolved COD constitutes about 70% of the total COD from their secondary treatment plant.

A loading of 0.87 corresponds to a dosage of

$$\frac{8.33 \times (60-7)}{0.87} = 507 \text{ lb carbon}/10^6 \text{ gal.}$$

This dosage is considerably higher than the dosage found at Pomona (250 lb/10^6 gal.), or at Lake Tahoe (300 lb/10^6 gal.)[4], but dosage is a concentration dependent term. The design removal (53 mg/l) is higher due to a higher influent concentration than the average removal at Pomona (37 mg/l) or at Lake Tahoe (10-35 mg/l). This higher removal accounts for the increase in the calculated dosage.

For the sake of completeness, the average loading for Lake Tahoe over 1-1/2 years onstream was about 0.25 lb COD/lb carbon as determined by planimeter measurements of their COD removal data. This relatively low loading reflects the lower influent concentration at Lake Tahoe and the fact that the carbon was sometimes taken out of the contactor before breakthrough due to plugging problems. Because the secondary effluent at Lake Tahoe was filtered to remove suspended matter prior to adsorption and the Pomona secondary effluent was not prefiltered, the Pomona capacity is higher since removal by filtration was included in the capacity. At Pomona, removal by filtration represents about 30% of the total capacity. Also, the rates of biological action may be lower at Lake Tahoe due to the colder climate.

It should be noted at this point that dosage is an indefinite term which should not be used without a knowledge of the concentration and removal of pollutant in the waste water. Loading (lb pollutant removed/lb carbon) is a more precise, meaningful term which has included removal in its calculation.

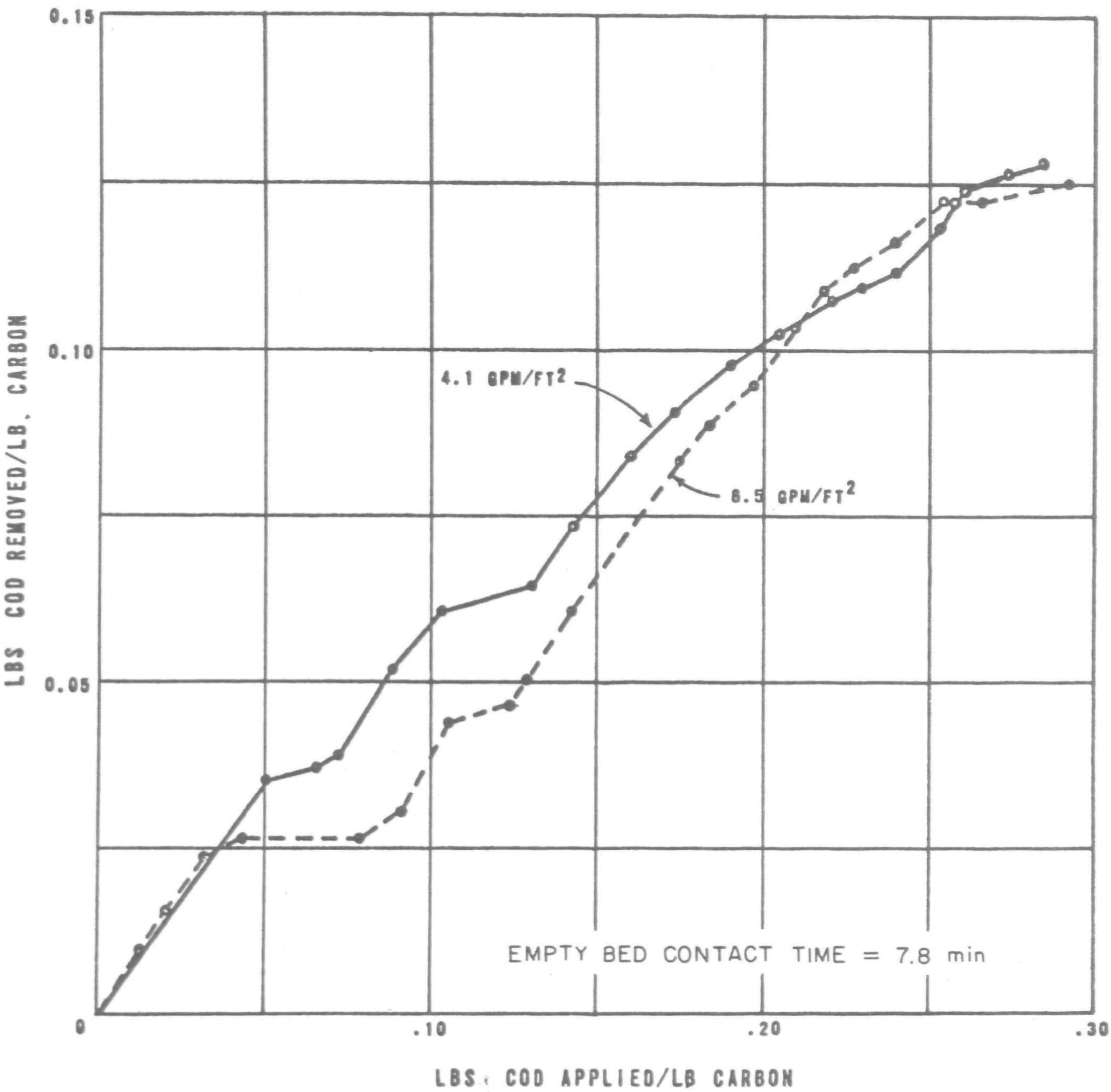
EFFECT OF LINEAR VELOCITY

With few exceptions, the literature on tertiary treatment is in agreement that linear velocity between 4 and 10 GPM/ft^2 has no effect upon the rate of adsorption. This is shown in several figures for various adsorbates:

Chemical Oxygen Demand (COD):	Figure 3
Alkyl Benzene Sulfonate (ABS):	Figures 4 to 7
Total Organic Carbon (TOC):	Figures 8 and 9
Color:	Figure 10

When examining velocity data, care must be taken to separate the effects of velocity, contact time, volume of water treated and length of time onstream. When removal data for two different velocities are taken on vessels of equal diameter and at the same contact time, the comparison should be made at the same length of time onstream; that is, the comparison is made after the same amount of water (and pollutant) has been applied to the same amount of carbon (the pounds·COD applied per pound of carbon is the same). This means that there will be a different total amount of water treated since the total amount of carbon in each case is different. To be completely comparable, parallel trains would be added to the low velocity experiment so that the same amount of water would be treated in the same length of time by the same amount of carbon as in the high velocity experiment. Of course, the percent removal accomplished by the parallel trains at the low velocity would not be any different from the removal accomplished by a single train. Therefore, the criteria for comparing velocity data should be the same amount of pollutant applied per pound of carbon at the same contact time rather than constant volume of water treated. The only case where data at constant total volume of water applied can be used is when the same amount of carbon is used for both velocities. This occurs when the cross sectional area for the low velocity case is larger by the ratio of velocities.

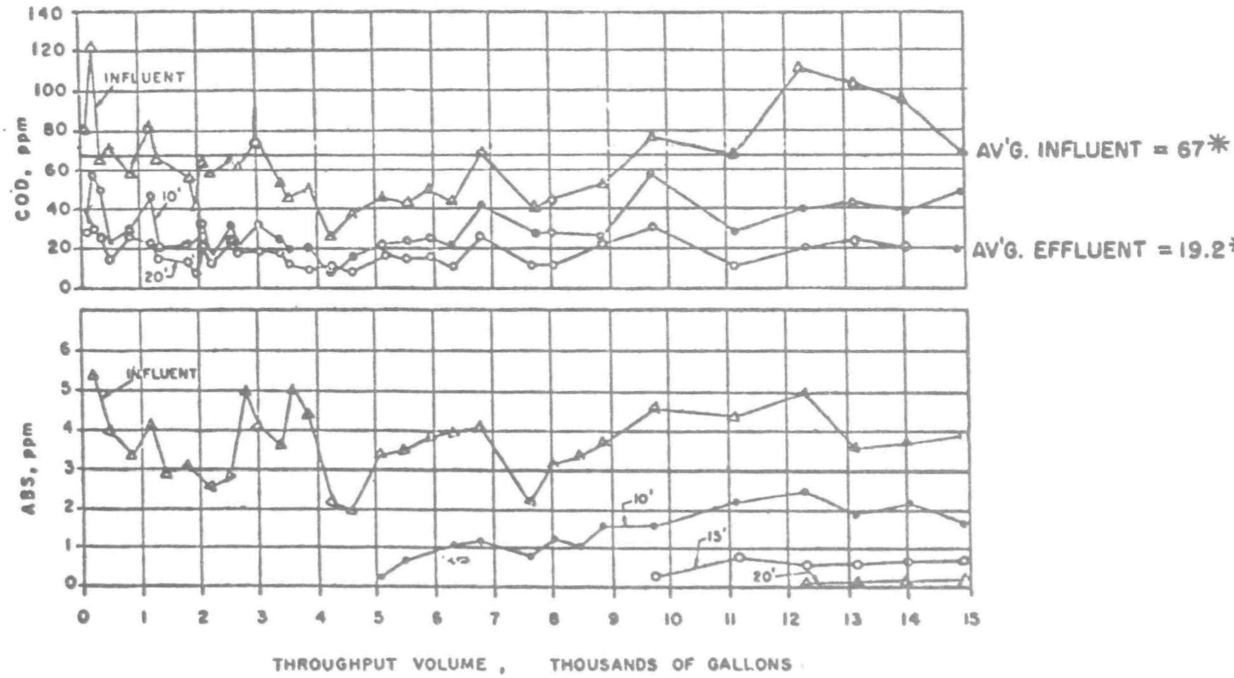
There are data which show that there is a contact time above which no more adsorption is accomplished in the column. Figures 6 and 8 show this effect at contact times greater than 15-20 minutes for virgin carbon adsorbing ABS (alkyl benzene sulfonate) and TOC (total organic carbon), respectively. The data in Figures 4 and 5 also fall in this range of contact time. Below this critical contact time, although there is an effect of contact time on removal, there is still no effect of velocity as seen in Figures 6-8. Figure 9 shows that for a contact time of about 3.5 minutes, the percent TOC removed is constant at 81% for velocities of 4, 7 and 10 GPM/ft^2.



REFERENCE 5

FIGURE 3

EFFECT OF VELOCITY ON CAPACITY
OF 8 X 30 MESH ACTIVATED CARBON FOR COD



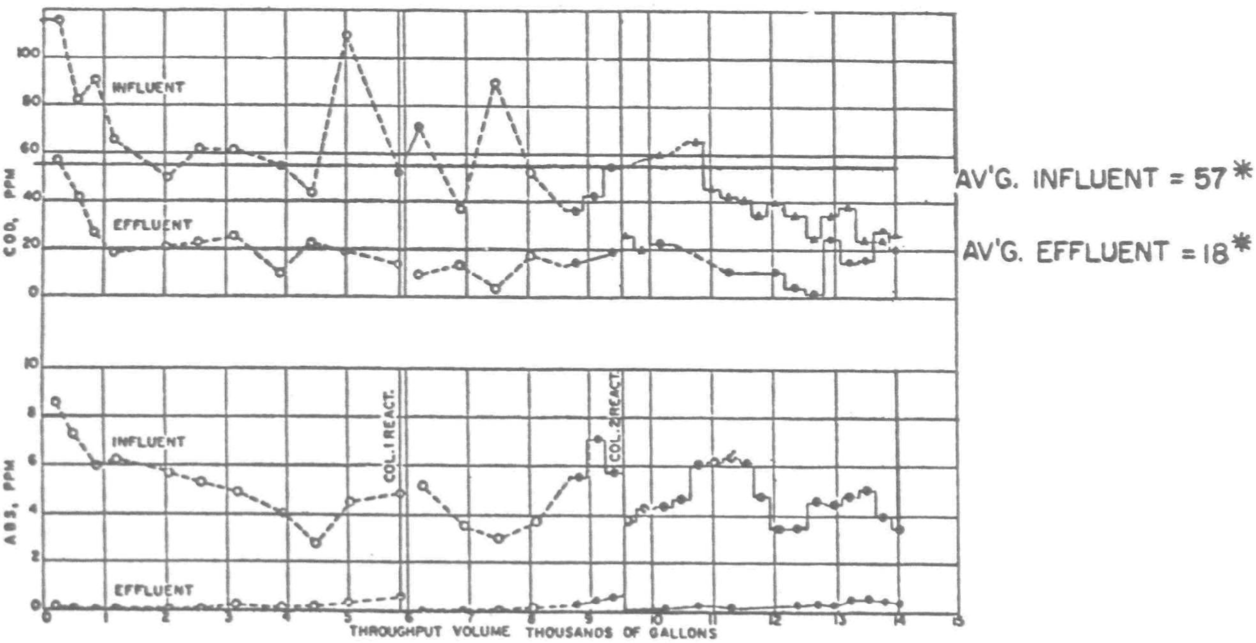
REFERENCE 6

FIGURE 4

COD BREAKTHROUGH CURVE AT 4 GPM/FT^2

REPRINTED BY PERMISSION, Joyce, R.S., *et al.*, *JWPCF*, 38, 816 (1966).

*ADDITIONS BY AUTHOR OF THIS REPORT.



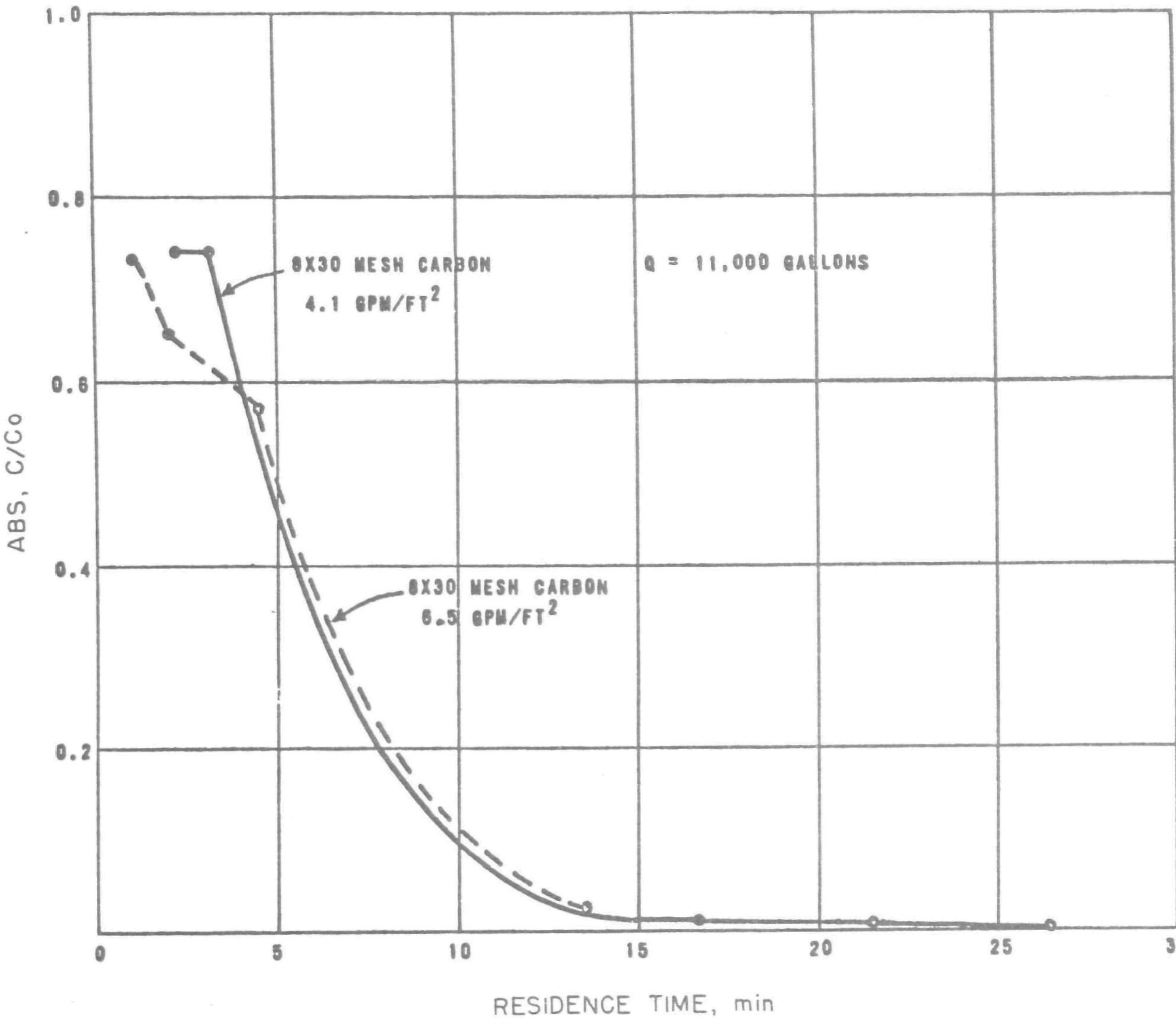
REFERENCE 7

FIGURE 5

COD BREAKTHROUGH CURVE AT 10 GPM/FT^2

REPRINTED BY PERMISSION, Joyce, R.S., *et al.*, *JWPCF*, 38, 815 (1966).

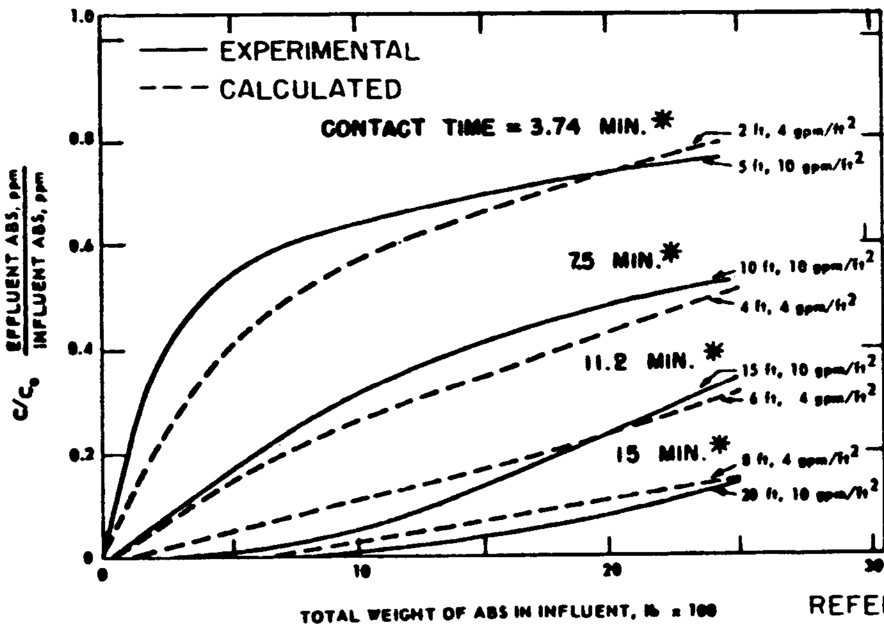
*ADDITIONS BY AUTHOR OF THIS REPORT.



REFERENCE 8

<u>FIGURE 6</u>

EFFECT OF VELOCITY AND CONTACT TIME
ON
ABS REMOVAL EFFICIENCY



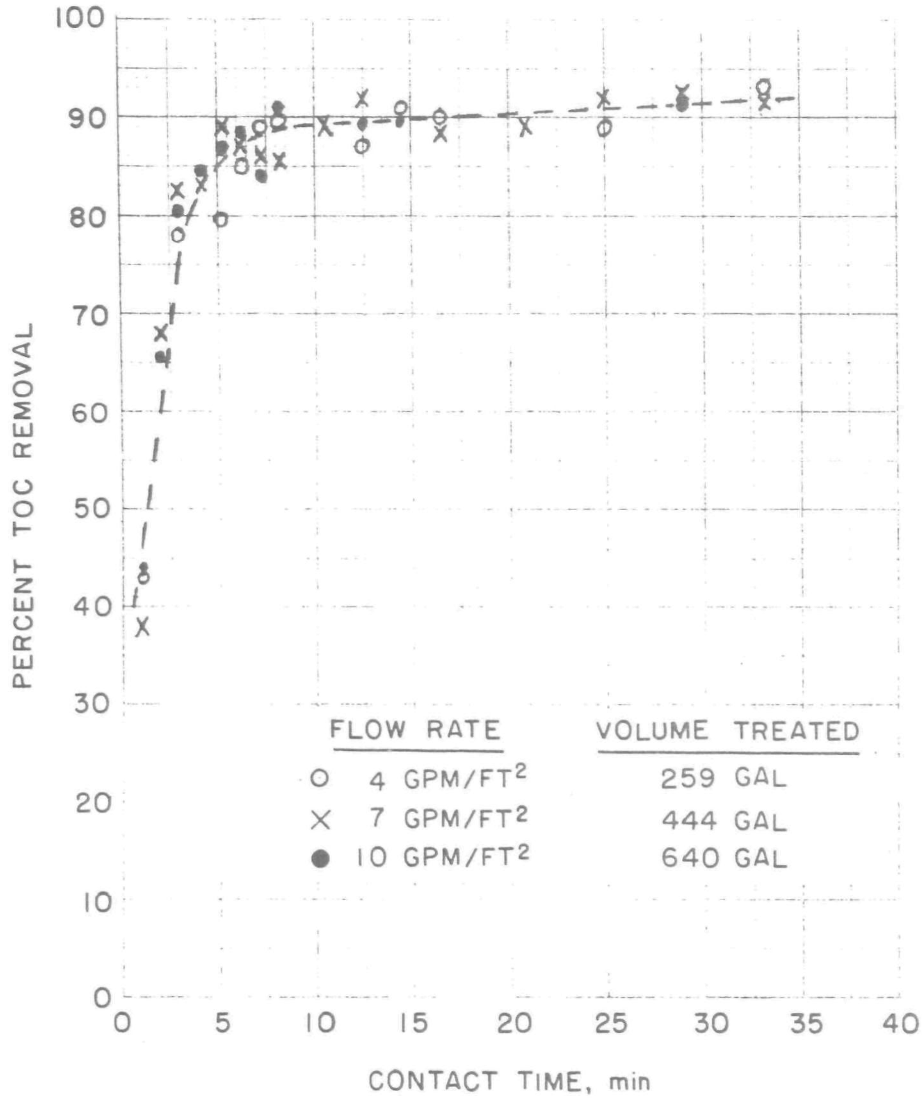
REFERENCE 9

AFTER TREATING EQUAL VOLUME OF WATER*

* ADDITIONS BY AUTHOR OF THIS REPORT

FIGURE 7

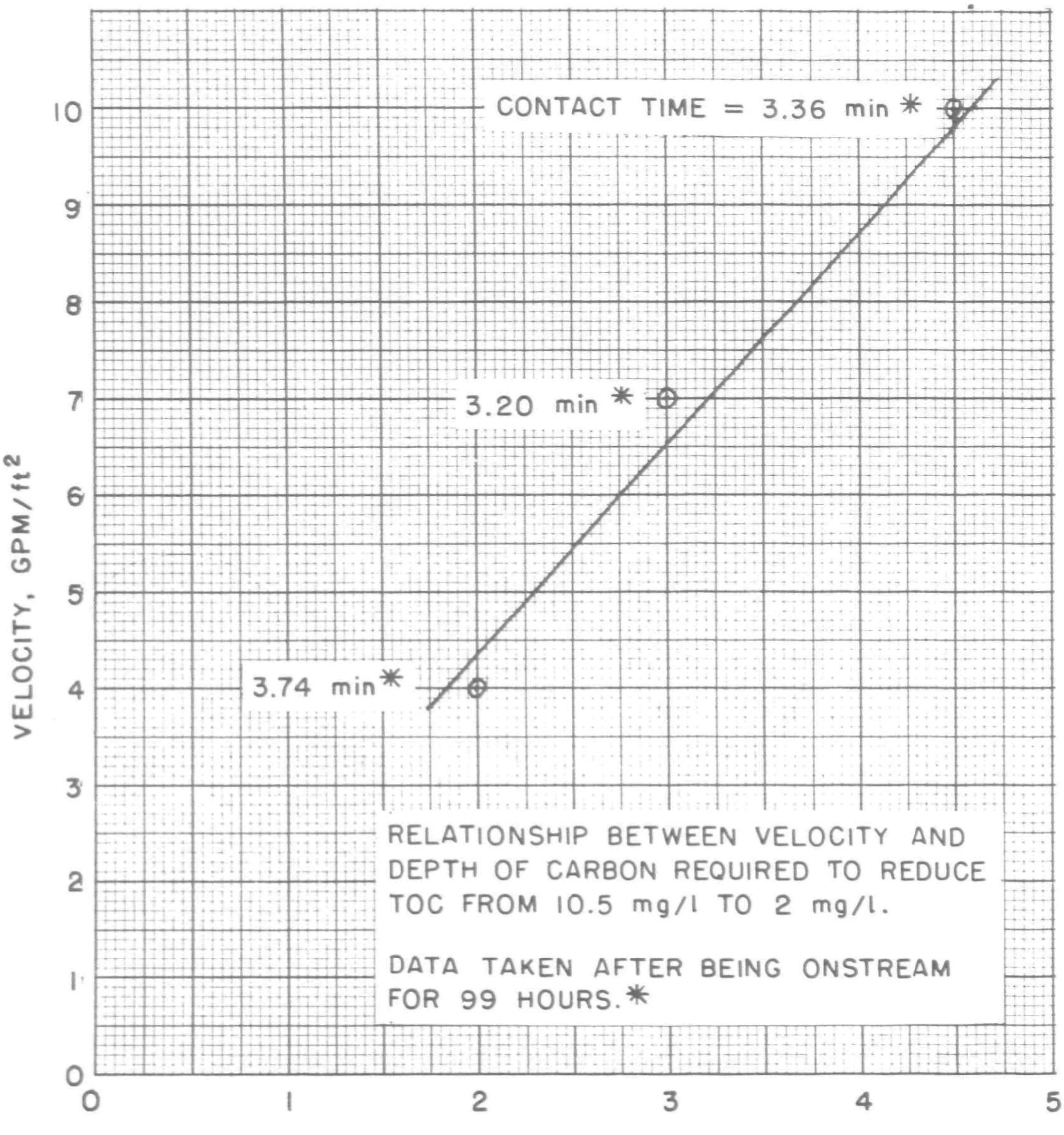
EFFECT OF VELOCITY AND CONTACT TIME
ON
ABS REMOVAL EFFICIENCY



REFERENCE 10

FIGURE 8

EFFECT OF VELOCITY AND CONTACT TIME
ON
TOC REMOVAL EFFICIENCY



REFERENCE 11

FIGURE 9
EFFECT OF VELOCITY ON TOC ADSORPTION

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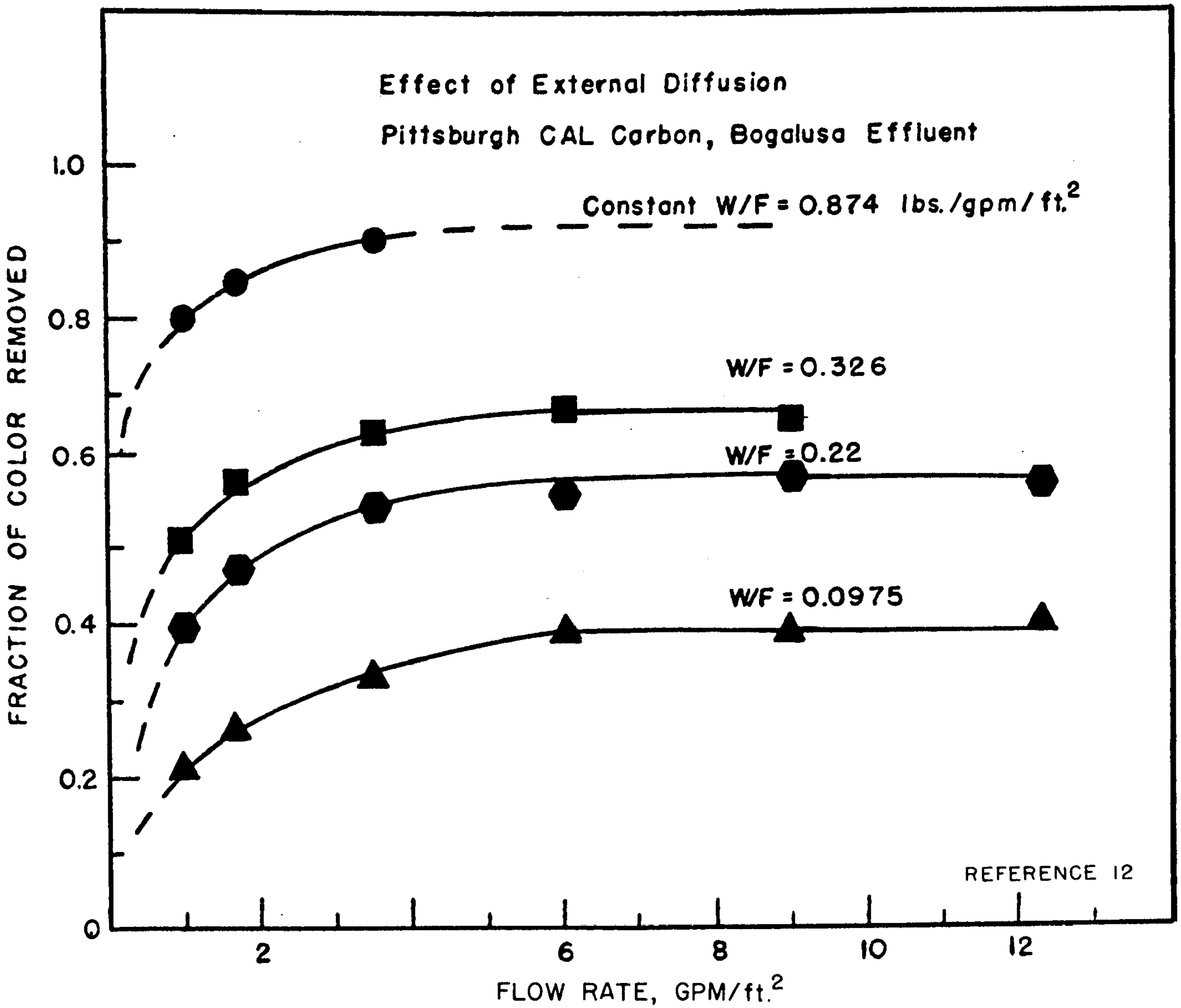

FIGURE 10

EFFECT OF VELOCITY ON COLOR REMOVAL

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Other evidence showing an effect of velocity can be seen in the following figures:

COD: Figures 4, 5, 12
ABS: Figure 13

In Figures 4 and 5 (COD removal), the fraction COD remaining (C/Co) is about the same for the curves at 4 and 10 GPM/ft^2, but the contact time varied as follows:

Velocity, GPM/ft^2	4	10
Volume treated, gal.	5,460	14,100
C/Co	0.298	0.316
Contact Time, Min.	37.4	14.9

When the C/Co data for COD removal shown in Figure 11 were recalculated at equal onstream times and replotted against contact time in Figure 12, a considerable effect of velocity was found. As is evident in Figure 12, the high velocity (10 GPM/ft^2) improves the removal of COD at the same contact time. Thus, the higher velocity will reduce the diameter of vessel required without appreciably increasing the depth of bed required. The advantage of high velocity may decrease, however, at long contact time. Unfortunately, the data represented in Figure 12 were the only data which could be found in a form suitable for analysis showing the effect of velocity on efficiency of removal for COD. Data should be obtained for adsorption of COD at high velocity at contact times up to about 40 to 50 minutes, or removal efficiencies up to 85 percent in order to complete the curve in Figure 12.

The data on ABS removal (Figure 13) are seen to be inconsistent with the data of other investigators (Figures 4 to 7). Again the higher velocity accomplishes a higher removal at the same contact time. The two curves do appear to be converging at long contact times at which point there is no effect of velocity, i.e., the same amount of carbon is required at both velocities in order to get the desired removal.

It is generally agreed that there is a velocity below which diffusion to the surface of the carbon may be controlling the rate of adsorption[15]; however, above this velocity, the resistance to diffusion to the surface of the solid adsorbent is reduced and the controlling rate step becomes one of surface adsorption or diffusion within the pores of the particle. In summary, few data on COD removal are available for the following conditions:

1. Velocities below 4 GPM/ft^2 where gravity flow systems would operate.

2. High velocities at long contact time.

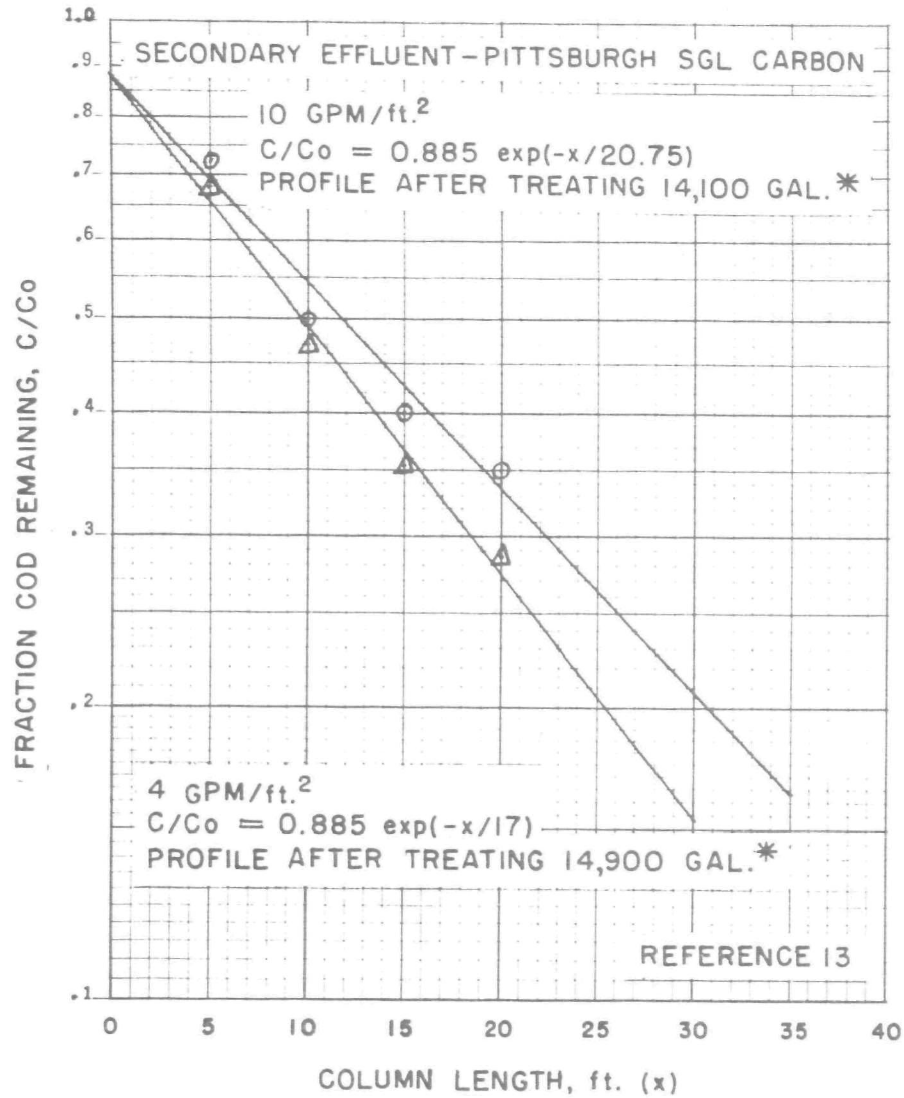

FIGURE 11

EFFECT OF VELOCITY ON COD REMOVAL EFFICIENCY

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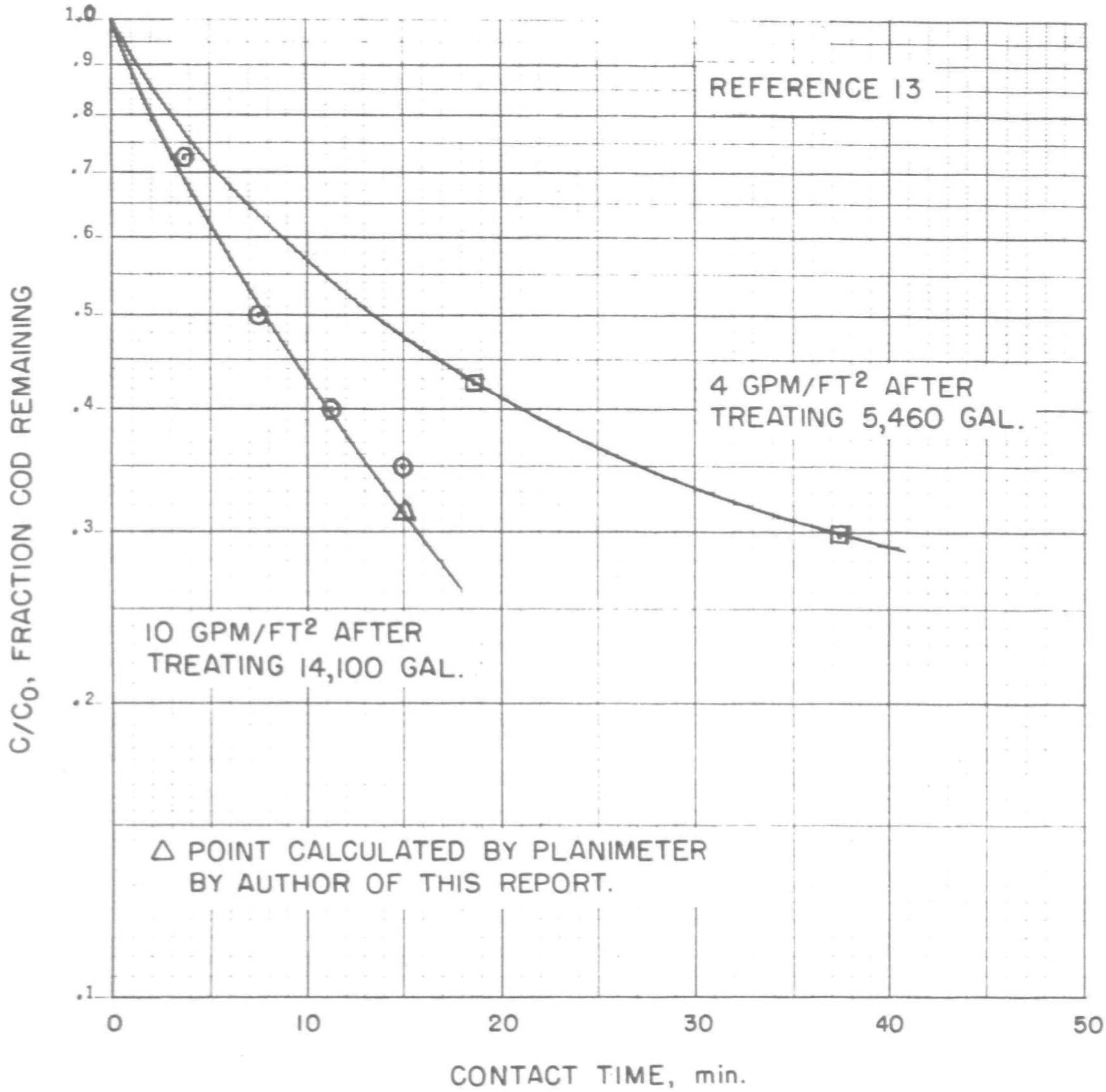
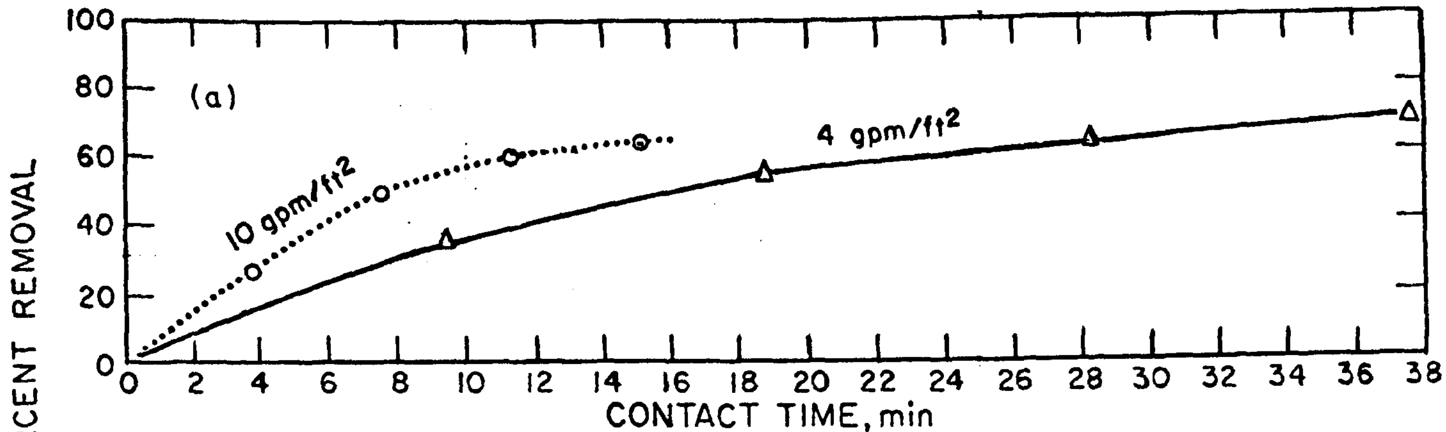
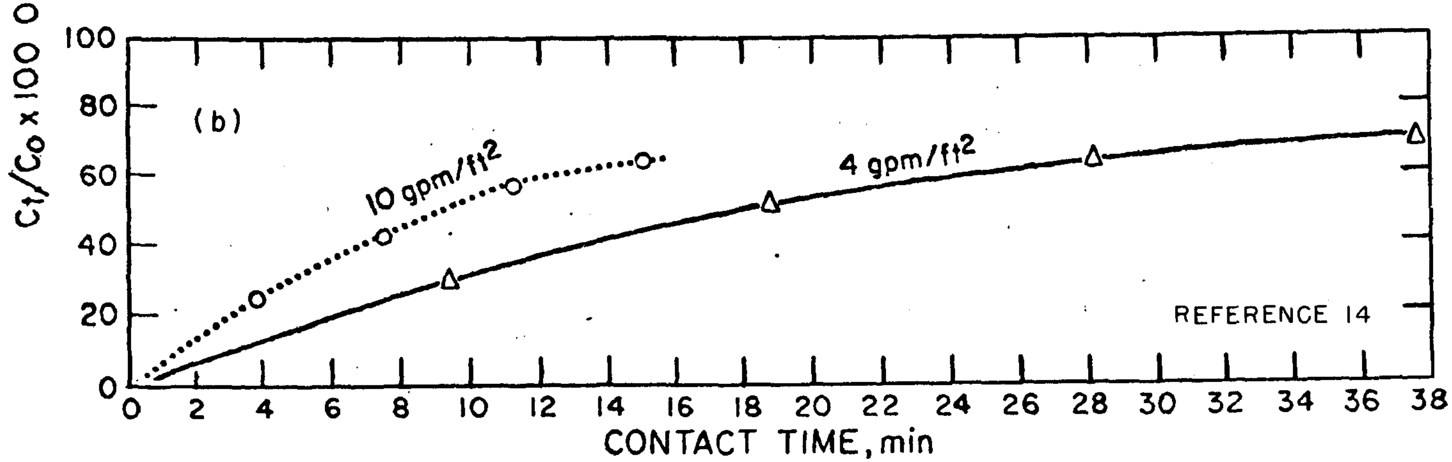


FIGURE 12

EFFECT OF VELOCITY AND CONTACT TIME
ON
COD REMOVAL EFFICIENCY



AVERAGE REMOVAL OVER PERIOD TO POINT OF YIELD OF 1,470 GALLONS OF WATER PER POUND OF CARBON IN SYSTEM



AVERAGE REMOVAL OVER PERIOD TO POINT OF YIELD OF 3,510 GALLONS OF WATER PER POUND OF CARBON IN SYSTEM

AVERAGE-REMOVAL CONTACT TIME RELATIONSHIP, FEASIBILITY STUDIES.

FIGURE 13

EFFECT OF VELOCITY AND CONTACT TIME ON ABS REMOVAL EFFICIENCY

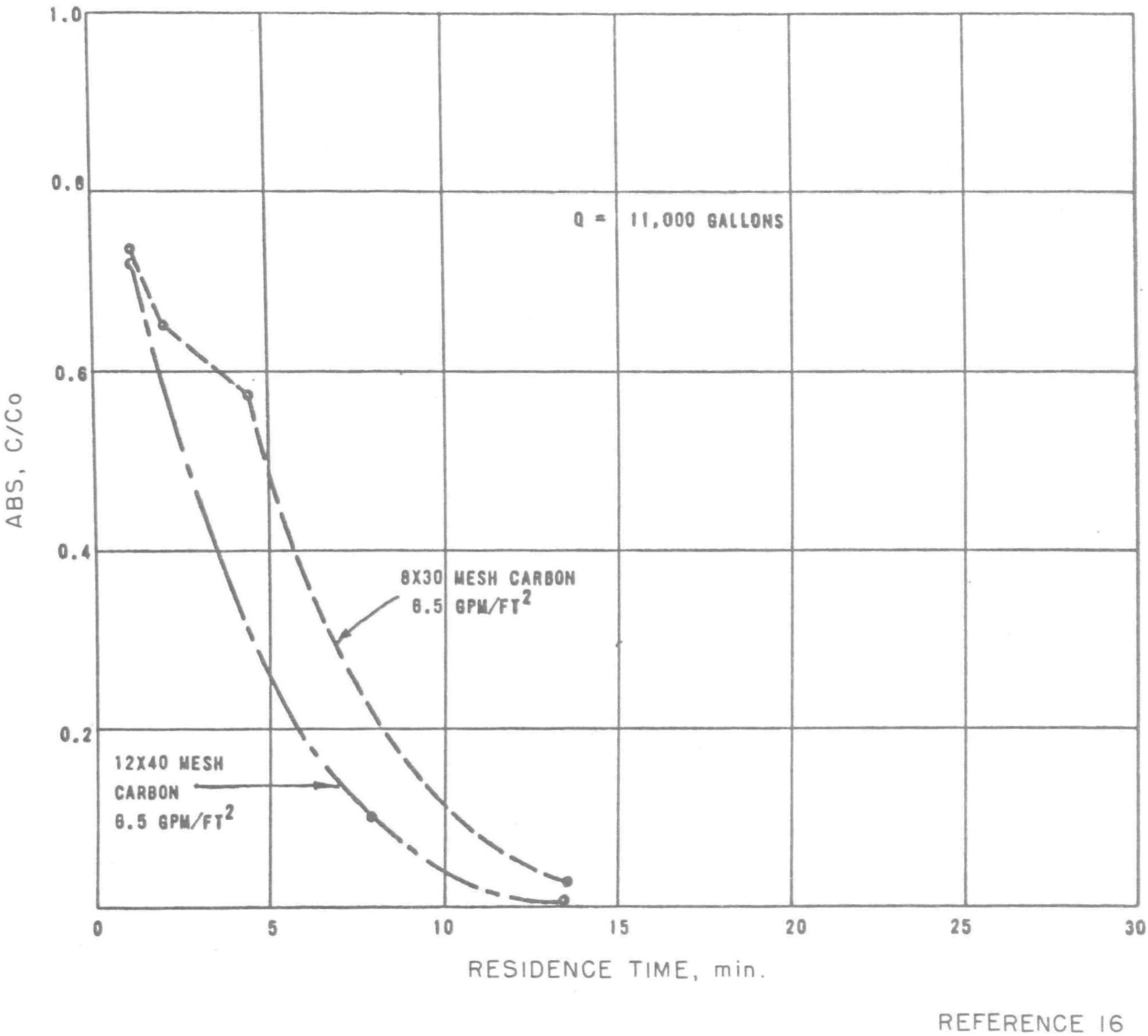
EFFECT OF PARTICLE SIZE

Since, theoretically, carbon particle size primarily affects the rate of adsorption and not the capacity of the carbon, adsorption columns operating close to saturation will be affected less by different particle sizes than will columns operating far from saturation. There is a complicating factor in some of these comparisons of particle size since the two sizes of Pittsburgh carbon do not have exactly the same pore size distributions. The effect of particle size is shown for ABS removal in data from Lake Tahoe (Figure 14). The small difference between the two carbons at both long and short contact times indicates that the carbons are approaching equilibrium for ABS. This is particularly evident at the long contact time since liquid concentration is approaching zero. Also at short contact time, the carbon is approaching equilibrium with the feed.

But so long as the carbon is far from equilibrium, an effect of particle size should be expected. When the data for COD removal (Figure 15) are plotted against contact time (Figure 16) in the same manner as Figure 14, an increasing difference between the two carbons can be seen for increasing contact time. This indicates that the carbon is still far from equilibrium for COD adsorption. There are other data which show less effect of particle size on COD removal (Figure 17).

When data for color removal (Figure 18) are replotted against particle size (Figure 19), it can be seen that since the lines of constant contact time seem to be converging with decreasing particle size, there is probably a particle size below which contact time will have no effect. However, this particle size appears to be less than 60 mesh which is below the range of commercial interest.

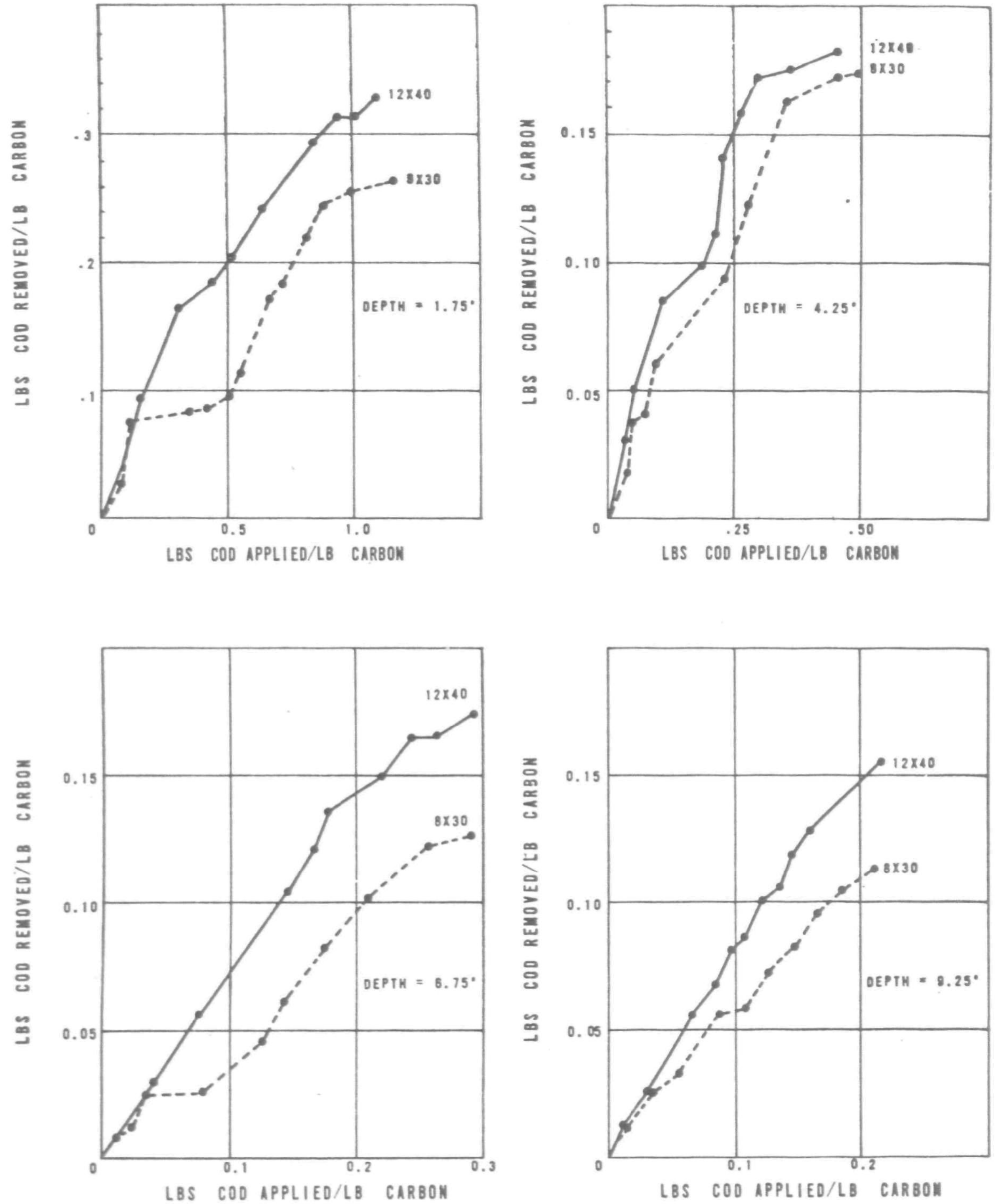
Since few data are available on the effect of particle size on COD removal, no precise quantitative recommendation can be made. But there probably will be a reduction on the order of 20 to 35% in allowable capacity in going from the 12 x 40 mesh carbon to 8 x 30 mesh at these relatively short contact times, based on the Lake Tahoe data (Figure 15). No data are available showing the effect of particle size after regeneration.



EFFECT OF CARBON PARTICLE SIZE ON ADSORPTION

FIGURE 14

EFFECT OF CARBON PARTICLE SIZE
ON
ABS REMOVAL EFFICIENCY



REFERENCE 17

FIGURE 15

EFFECT OF CARBON PARTICLE SIZE ON COD ADSORPTION

FLOW RATE = 6.5 GPM/FT^2 *

* ADDITION BY AUTHOR OF THIS REPORT

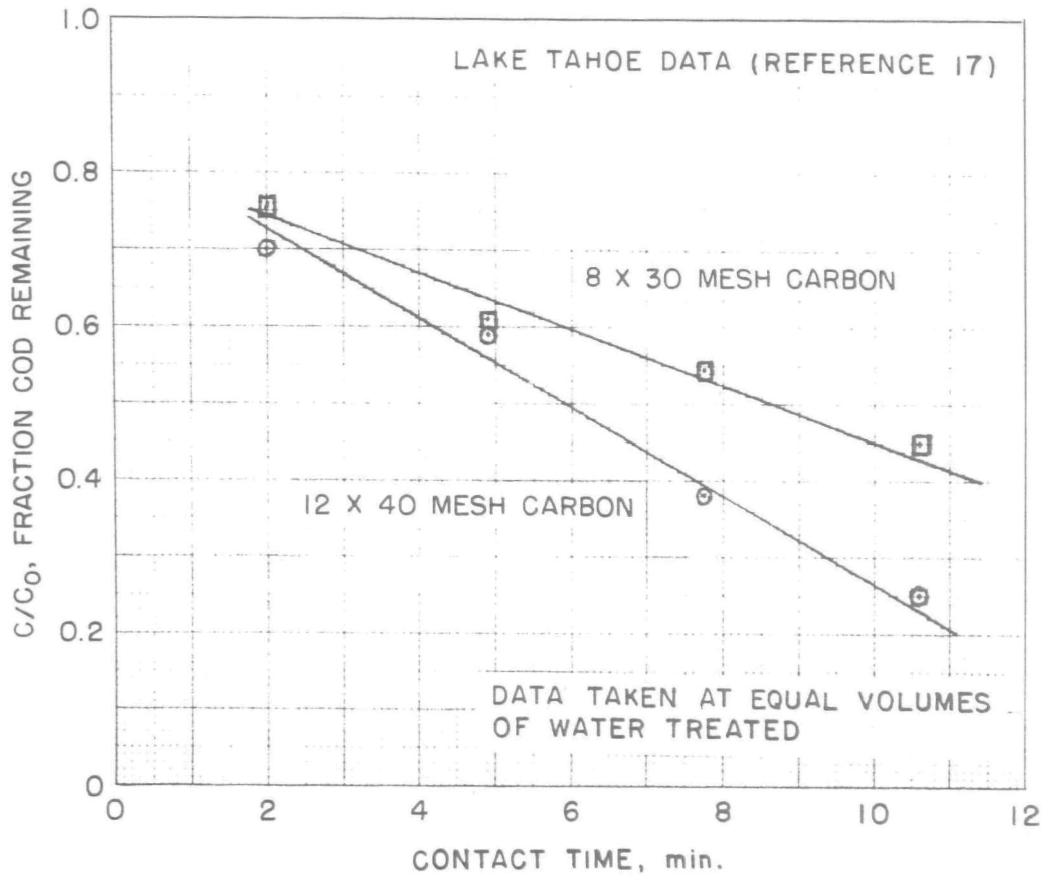
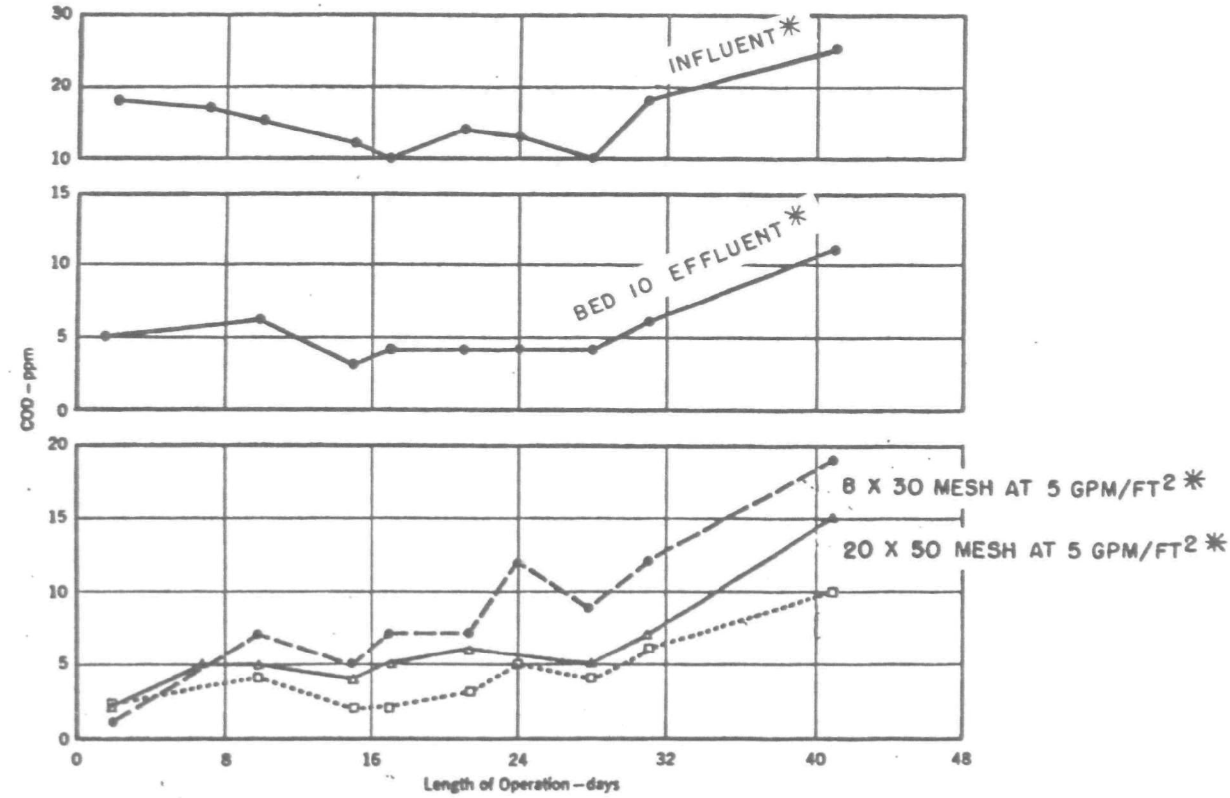


FIGURE 16

EFFECT OF CARBON PARTICLE SIZE
ON
COD REMOVAL EFFICIENCY



Passage of COD With Time, for Various Operating Conditions

REFERENCE 18

FIGURE 17
EFFECT OF CARBON PARTICLE SIZE ON COD ADSORPTION

REPRINTED BY PERMISSION, Dostal, K.A., *et al*, *JAWWA*, 57, 670 (1965)

*ADDITIONS BY AUTHOR OF THIS REPORT

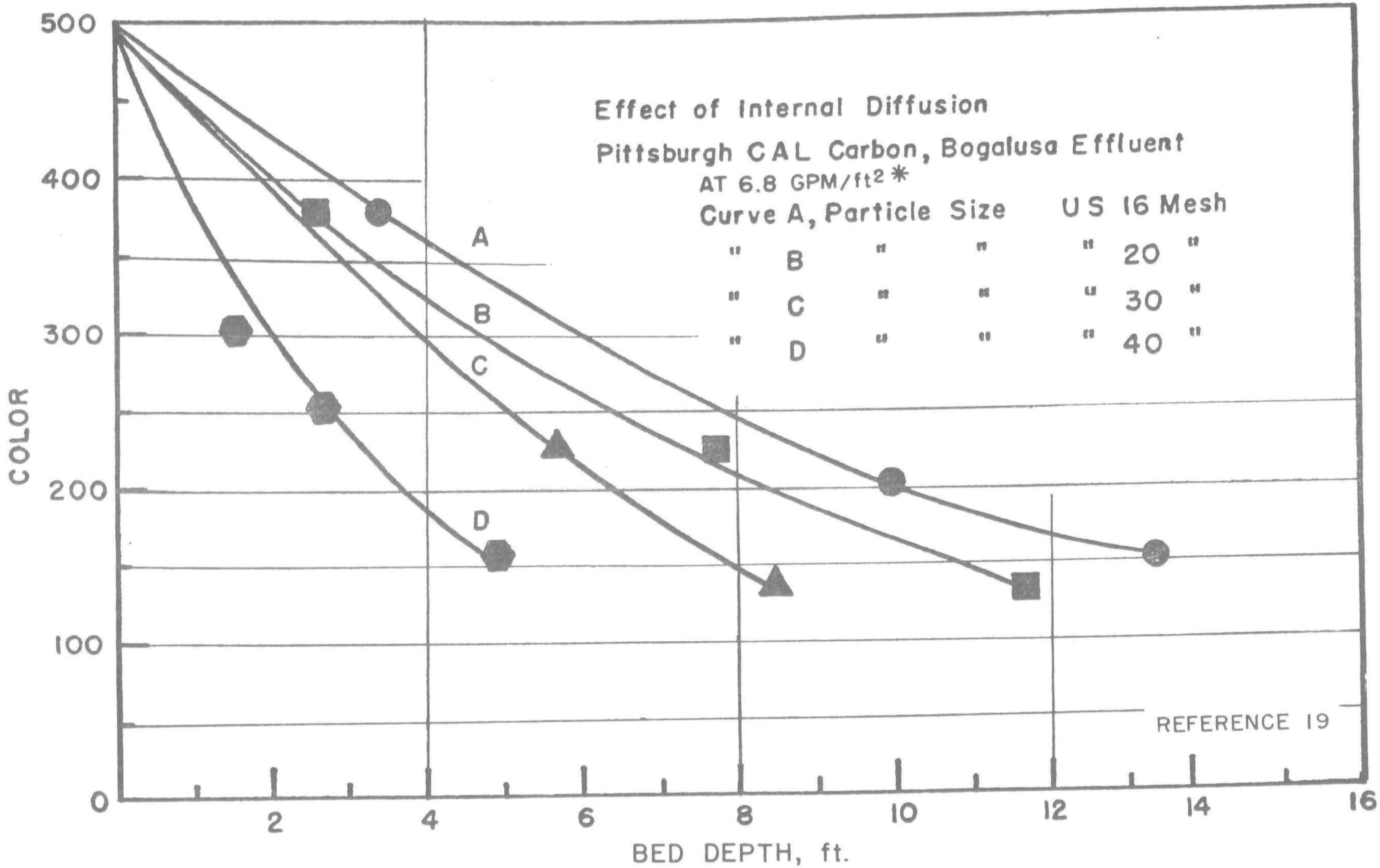


FIGURE 18

EFFECT OF CARBON PARTICLE SIZE ON COLOR REMOVAL

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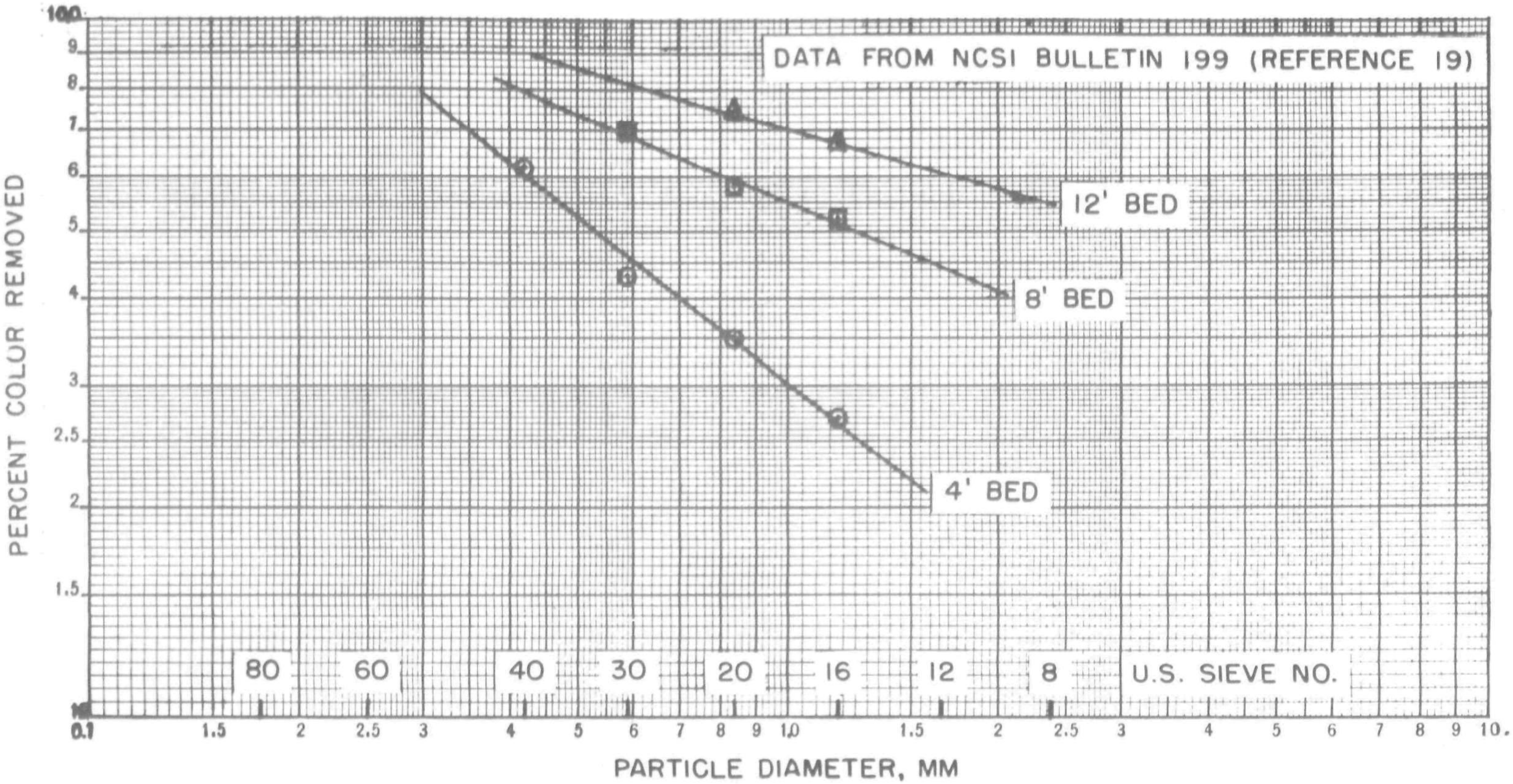


FIGURE 19

EFFECT OF CARBON PARTICLE SIZE
ON COLOR REMOVAL EFFICIENCY

EFFECT OF REGENERATION

Care must be taken when analyzing data for regenerated carbon to account for the effect of the make-up carbon which is usually added as virgin carbon. For example, when the regeneration losses are 10%, there will be little of the original carbon present after ten regenerations; that is, the original carbon will have been replaced by virgin make-up carbon. The assumption here is that the carbon which has been regenerated will be preferentially lost in subsequent regenerations, especially as the particles become smaller and weaker.

Losses in regeneration can occur by attrition in the pipeline and conveyors, by attrition, decrepitation, gasification and burning in the regeneration furnace, or by thermal shock in the quench tank. Variables which may affect the regeneration loss are particle size and shape, type and degree of loading, amount of handling, regeneration severity, and carbon base material (such as bituminous coal, wood, coconut shell, petroleum coke, etc.) There is a need to pin-point where carbon losses occur and, perhaps, to develop a rounder carbon particle which would be more attrition-resistant. Another major loss of carbon could be by gasification which could be as high as 50% of the carbon per hour.[20] Regeneration losses should be higher for the smaller size carbon since the smaller particle is more active for adsorption and, therefore, should be more active for burning and gasification. The comparison between adsorption and burning can be made as long as the regeneration temperature is low enough so that burning is controlled by pore diffusion and not by diffusion to the external surface. Under this condition, burning is a pore diffusion phenomenon as is adsorption. If the burning rate at this temperature were entirely an external surface phenomenon, then activated carbon would have the same reactivity as coke which is clearly not the case.[21] The difference in particle size may account for part of the difference between the 7% to 10% loss experienced at Pomona with 16 x 40 mesh carbon and the 5% loss at Lake Tahoe with 8 x 30 mesh carbon. Also, the operation of the regeneration furnace on a continuous vs. batch basis will affect the carbon loss. Carbon loss should be higher in batch operation than in continuous operation because the carbon left in the furnace between regenerations can be burned during subsequent shutdown and startup. Lake Tahoe has minimized this problem by placing sand on the hearths of the furnace to prevent the residence of carbon under the rabble arms.

Results of recent work done by the MSA Research Corporation [22] for the FWPCA show that the conditions used in the multiple-hearth regeneration furnace are far from the optimum with respect to regeneration gas flow rate, carbon residence time in the furnace, and carbon temperature. It was found that a regeneration loss of 2% could be obtained in a 3.25 in. I.D. x 65 in. long rotary tube regenerator with a lower gas flow rate, longer carbon residence time and higher carbon temperature. However, it has not been established whether or not all the carbon loss at Pomona and Lake Tahoe occurs in the regeneration furnace.

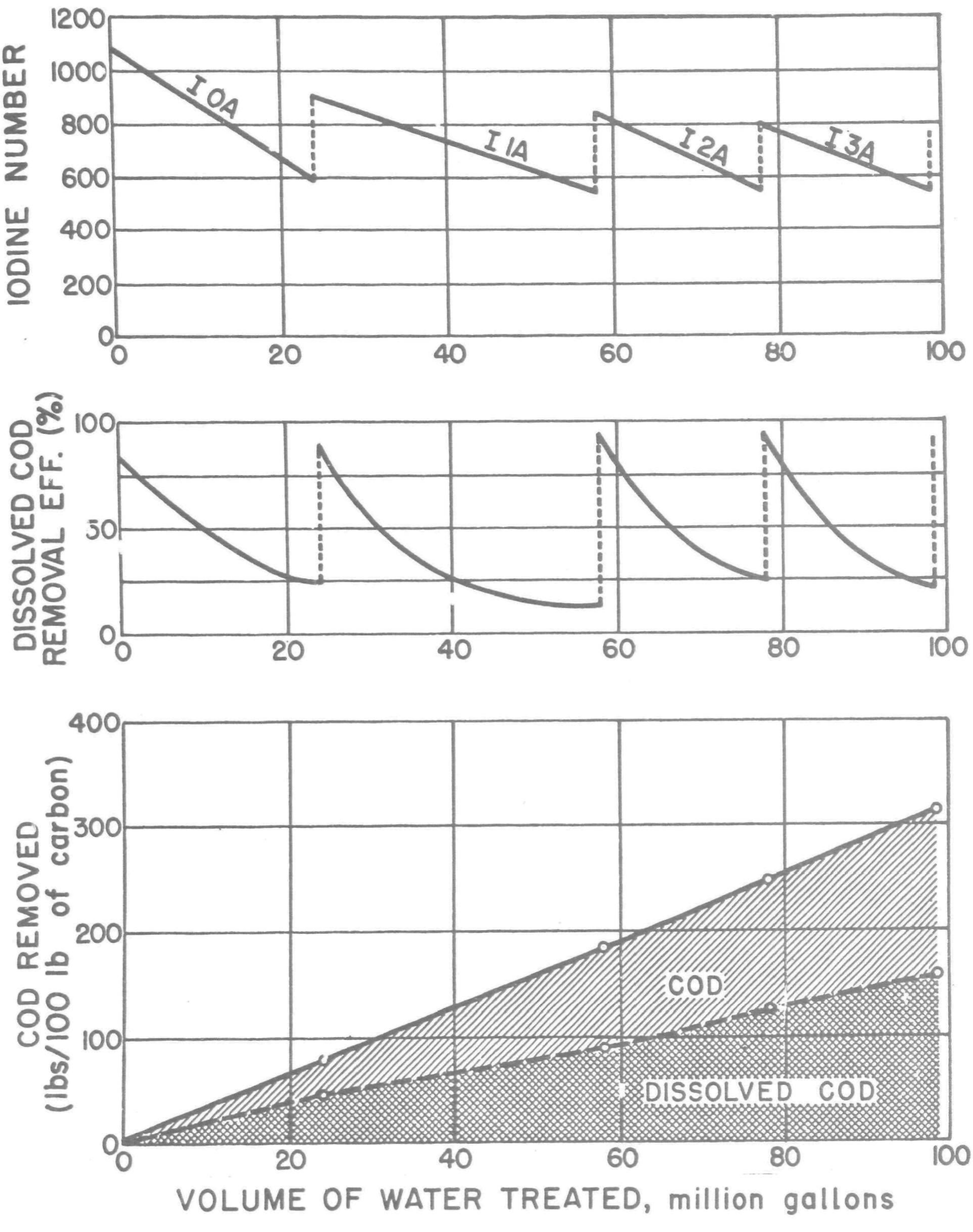
After each of three regenerations in the accelerated contactor at Pomona, the carbon has come back to 90% dissolved COD removal efficiency (Figure 20), despite a continuously decreasing iodine number. Furthermore, this decrease seems to occur during the first regeneration and not during any subsequent regeneration (Figure 21). However, the loadings for dissolved COD at breakthrough have dropped from 0.46 lb/lb to 0.29 lb/lb after seven regenerations (Figure 22). It should be noted here that the seven regenerations on the accelerated contactor represent about seven years operation in the main carbon column at Pomona. The capacity shown for seven regenerations should not be extrapolated since it still appears to be decreasing (Figure 23). Furthermore, extrapolation may be unwarranted, since an economic evaluation may reveal that the carbon should be replaced after several years use rather than accept a decrease in capacity.

The concentration profile for COD for a bench-scale column operating at 1.84 GPM/ft^2 is the same for virgin and sixth cycle regenerated carbon (Figure 24). The regeneration losses at Pomona have averaged about 10%.

Lake Tahoe shows an apparent 15% decrease in ABS removal after one regeneration in a test made with pure regenerated carbon (Figure 25). Lake Tahoe is apparently taking no economic debit for decreasing capacity with regeneration since they have seen no increase in required dosage in 1-1/2 years' operation.[4] The regeneration losses at Lake Tahoe average 5%.

The Pittsburgh Activated Carbon Company has found a 35% decrease in equilibrium capacity for ABS after 16 regenerations (Figure 26). Adsorption isotherms for COD on type SGL virgin and carbon regenerated 16 cycles do not appear to be significantly different (Figure 27). The authors later show, however, that the adsorption isotherm will be of no value in design since the apparent capacity of the carbon in an operating column is 50 to 100% greater than the isotherm capacity.[1] The authors assert that this increased capacity may be due to biological activity. Their regeneration losses averaged 4.2%. Another paper by the same authors shows the removal efficiency for COD after 16 regenerations (Figure 28). Here the difference between virgin and regenerated carbon does not appear until the carbon has treated a large volume of water (dosage 100 lb/10^6 gal.). This finding is consistent with the results of the bench-scale test with sixth cycle carbon at Pomona shown in Figure 24, indicating that one of two things may be occurring during regeneration:

1. There may be some degradation of the pore structure which results in slower pore diffusion which becomes the rate controlling step as the carbon becomes more heavily loaded.



REFERENCE 23

FIGURE 20

PERFORMANCE OF ACCELERATED CONTACTOR No. I

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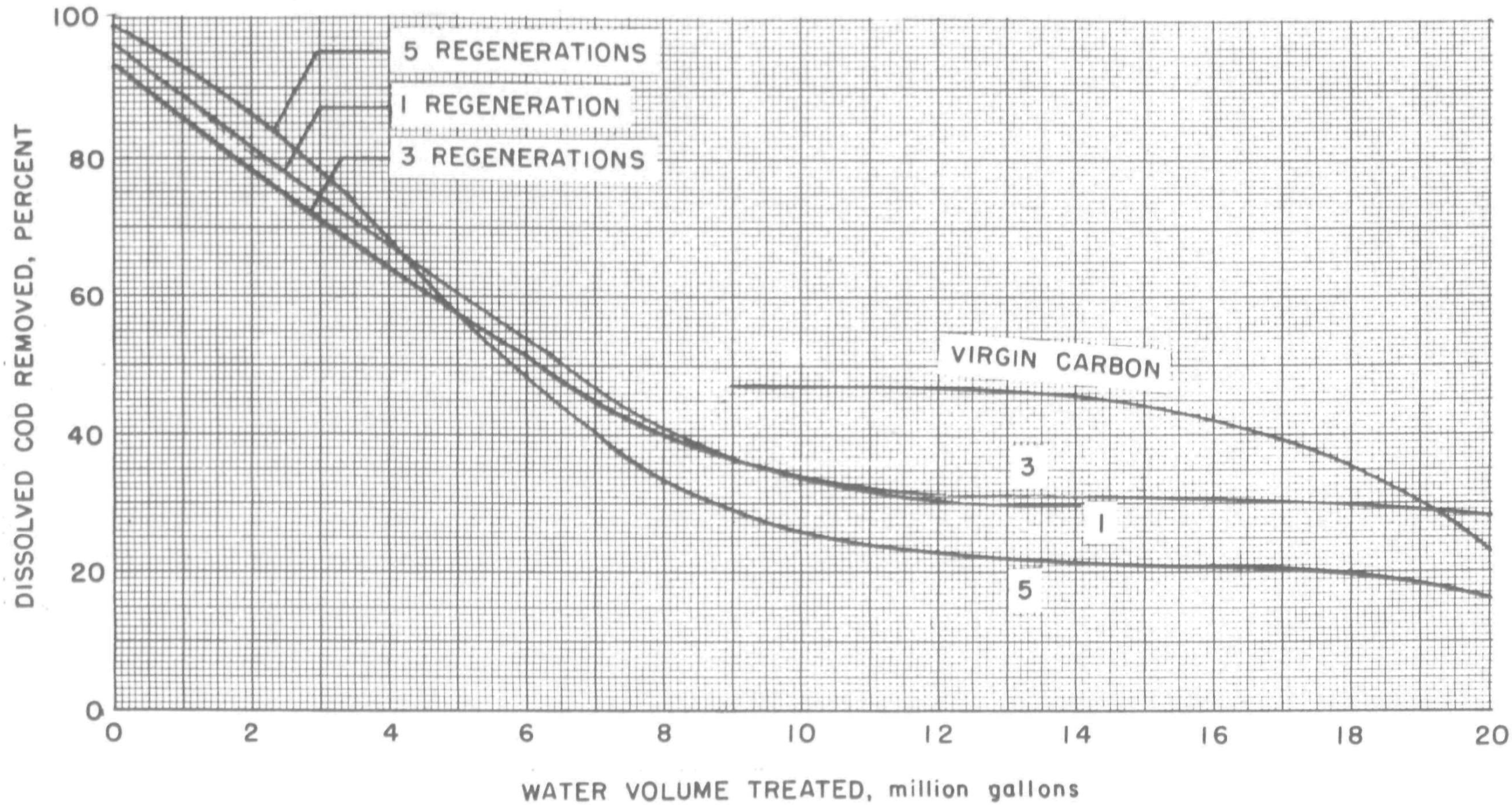
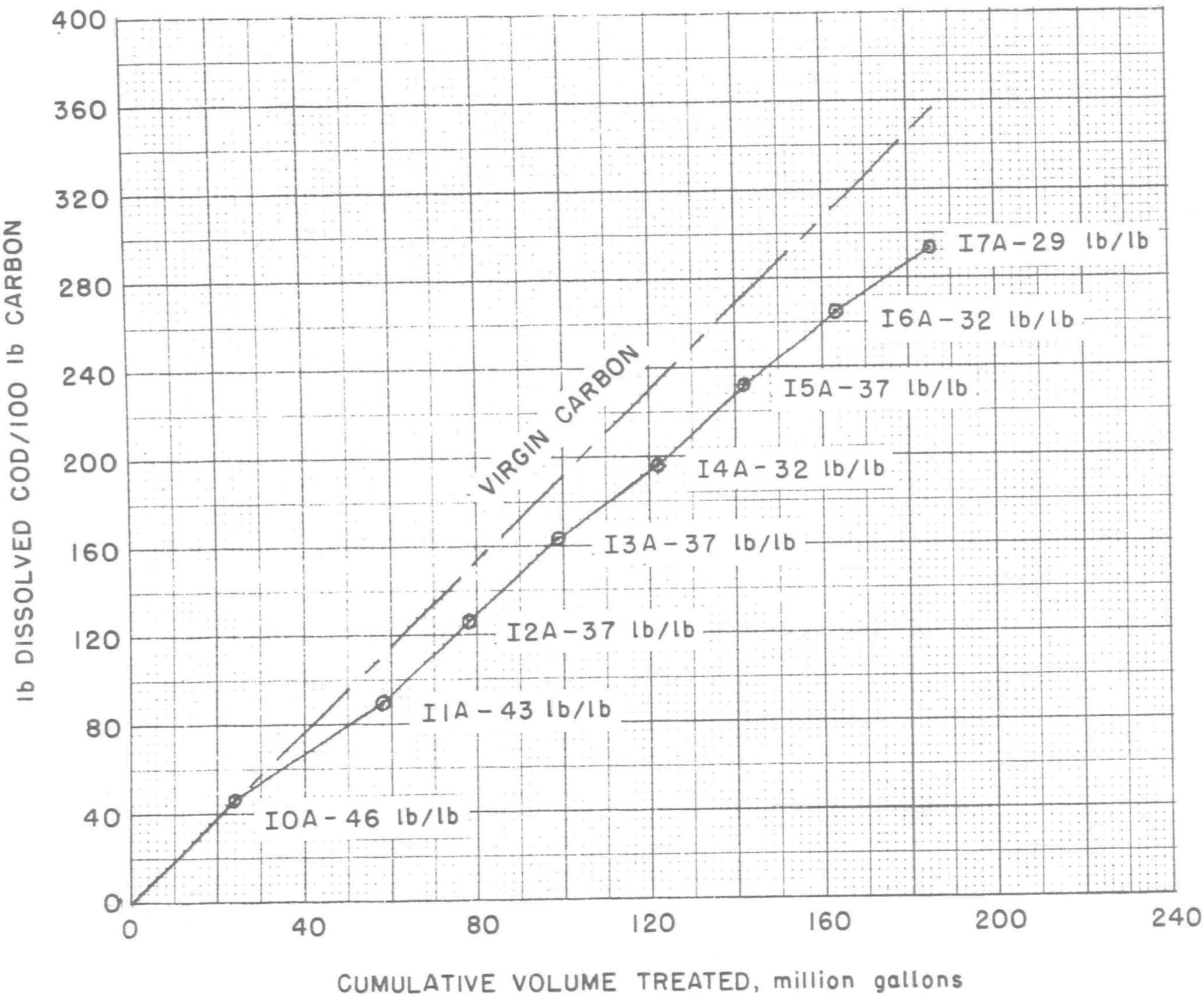


REFERENCE 24

FIGURE 21

EFFECT OF REGENERATION ON COD REMOVAL EFFICIENCY

*ADDITIONS BY AUTHOR OF THIS REPORT



REFERENCE 25

FIGURE 22

EFFECT OF REGENERATION ON CAPACITY FOR COD

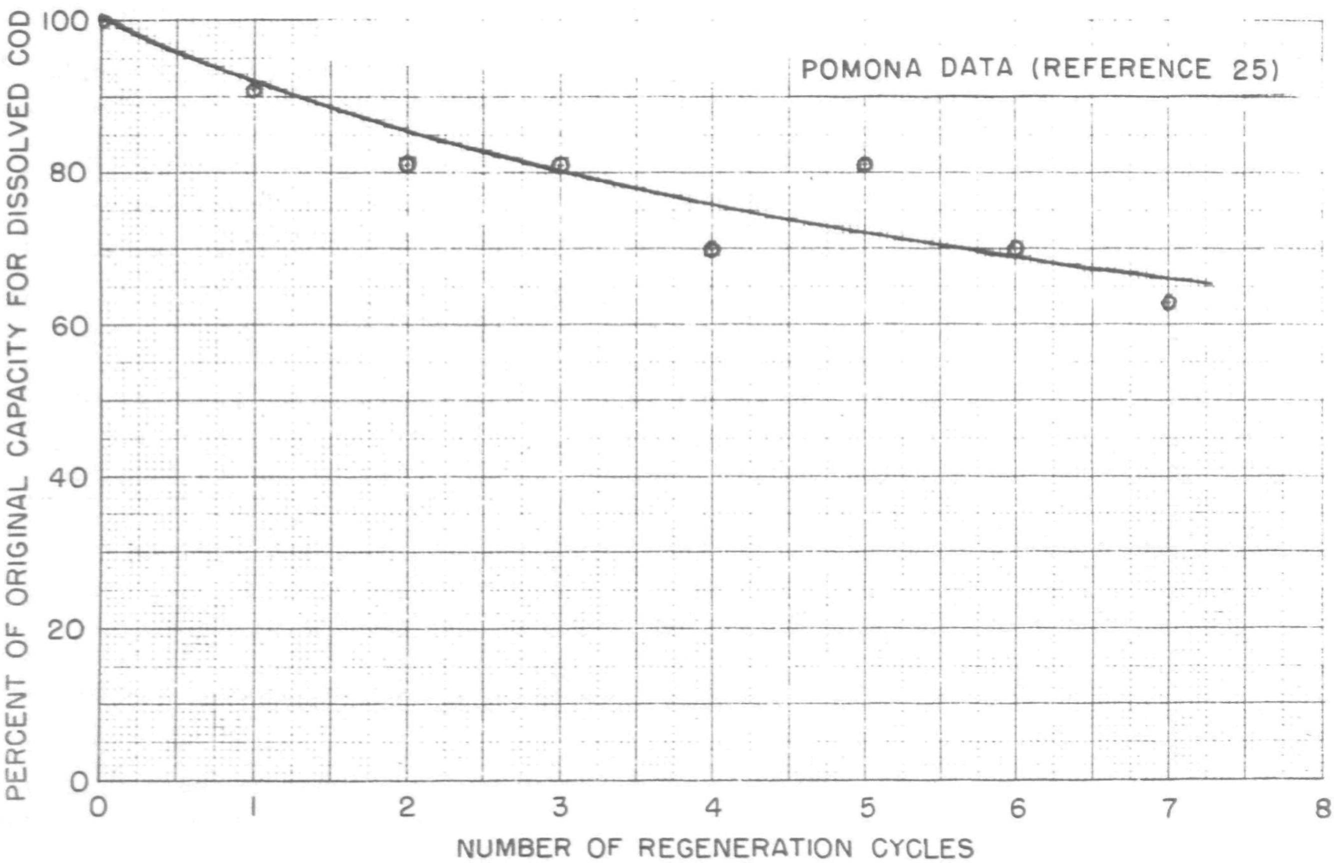


FIGURE 23
EFFECT OF REGENERATION
ON
CAPACITY FOR COD

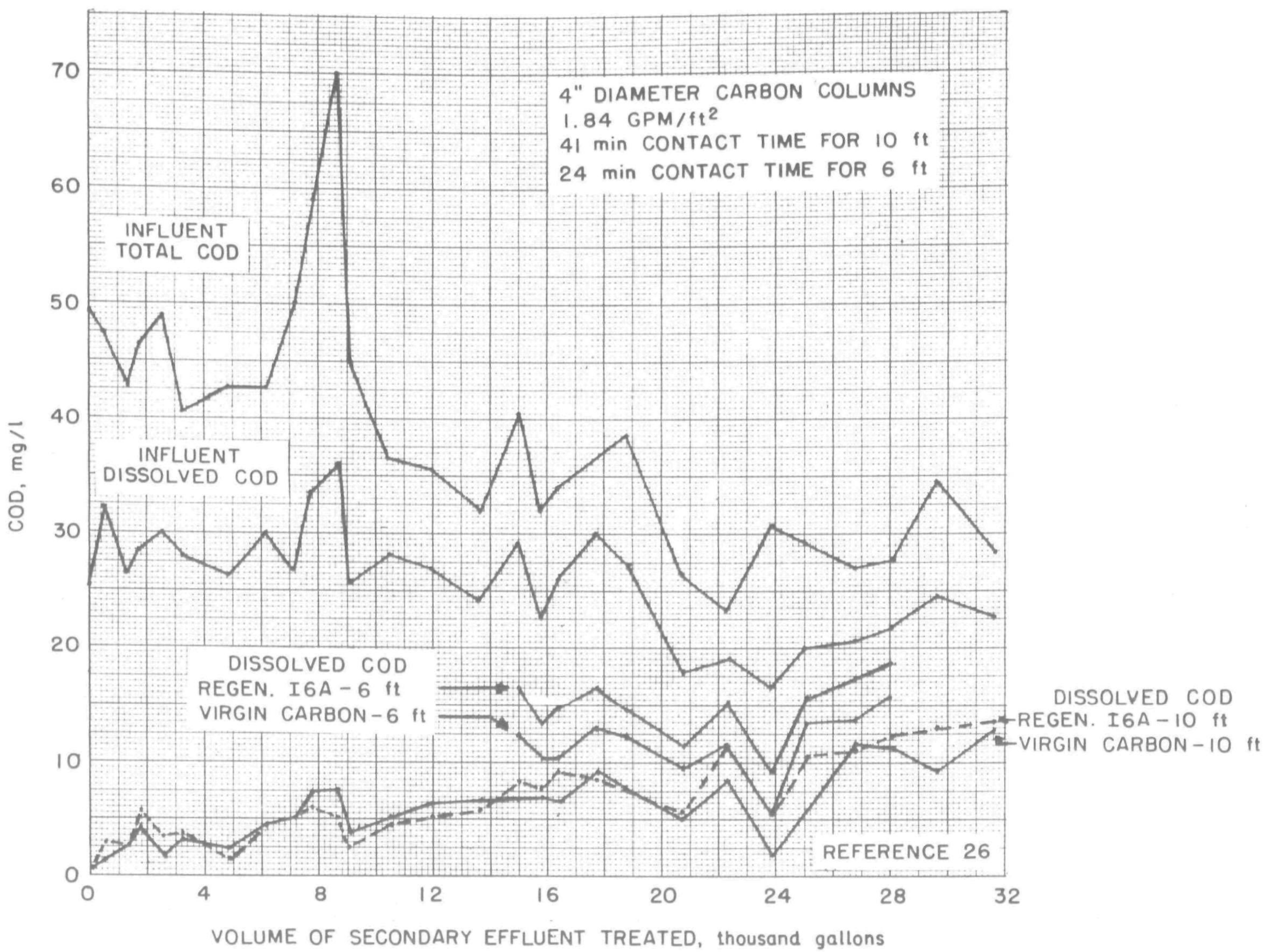
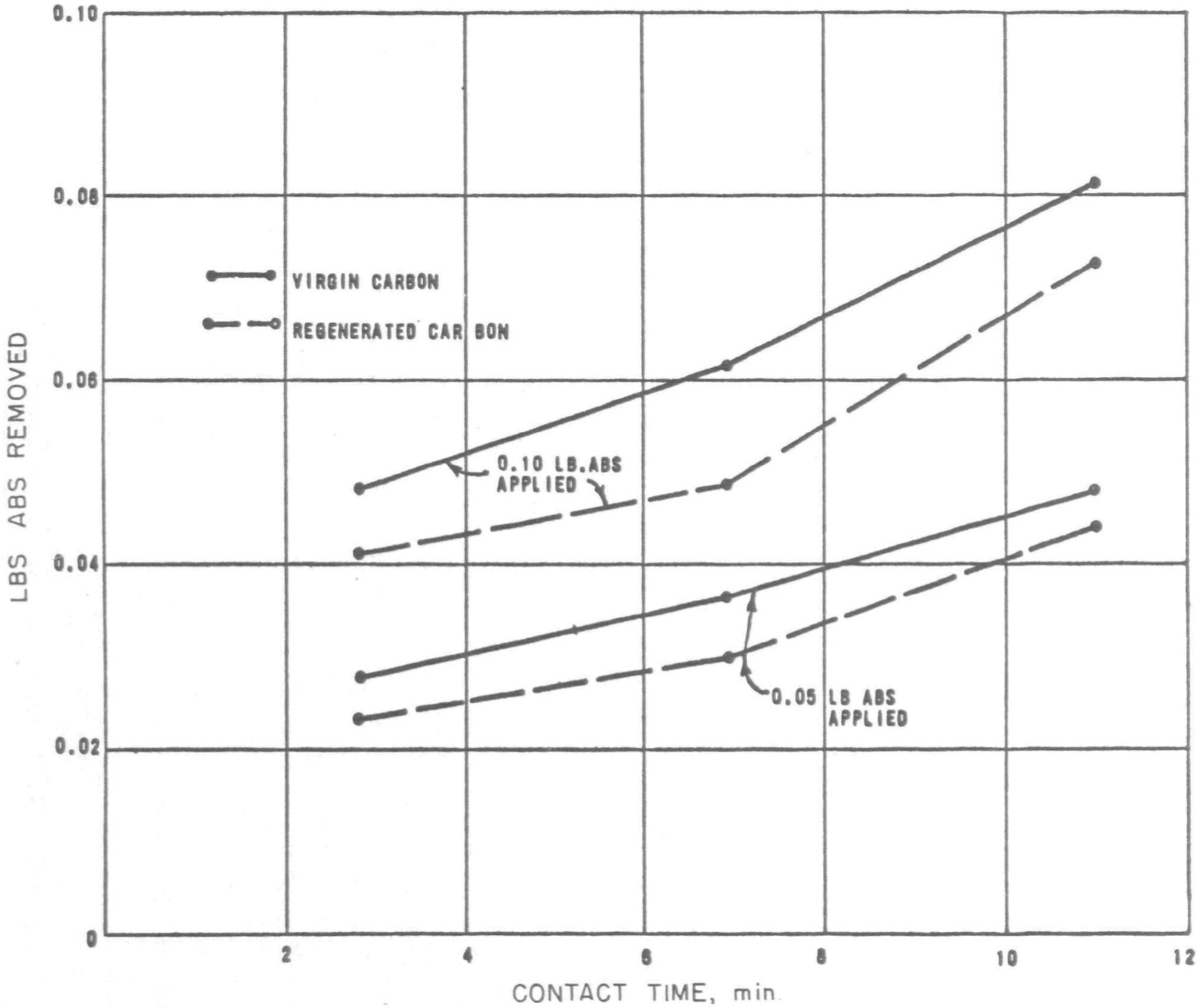


FIGURE 24

EFFECT OF REGENERATION ON COD ADSORPTION

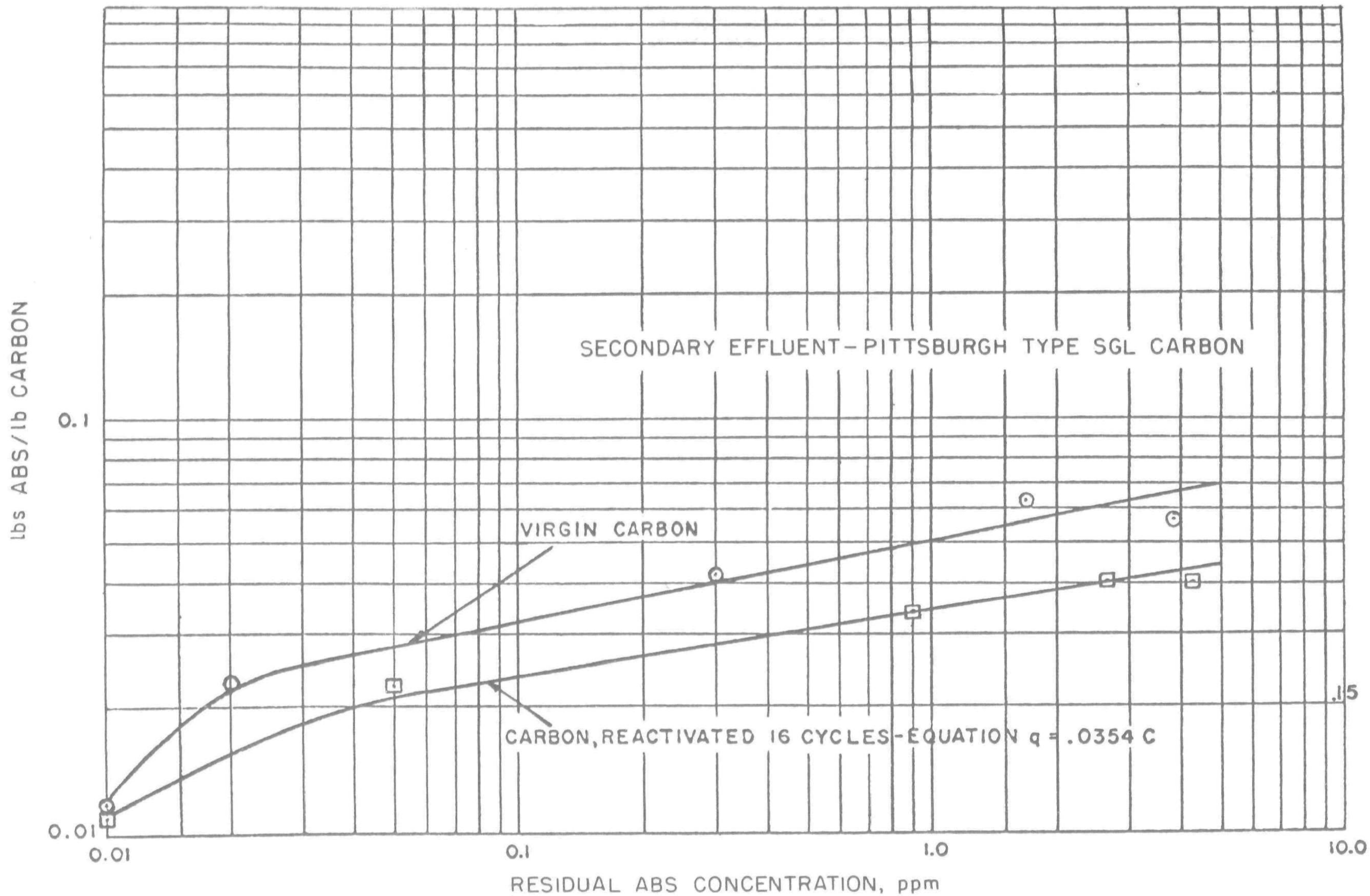


COMPARISON OF VIRGIN AND FIRST CYCLE REGENERATED CARBON FOR ADSORPTION OF ABS AS FUNCTION OF CONTACT TIME

REFERENCE 27

FIGURE 25

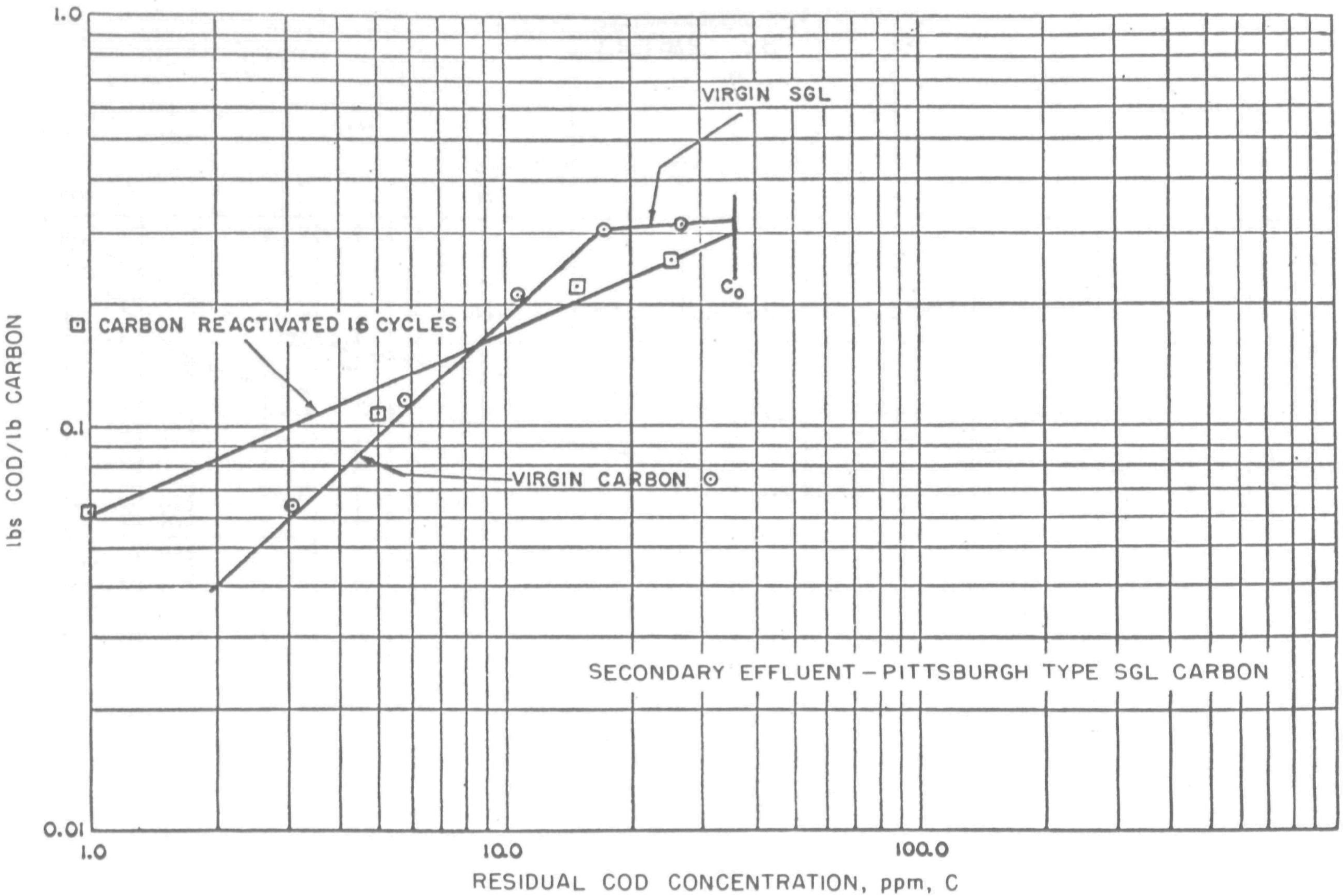
EFFECT OF REGENERATION ON ABS ADSORPTION



REFERENCE 28

FIGURE 26

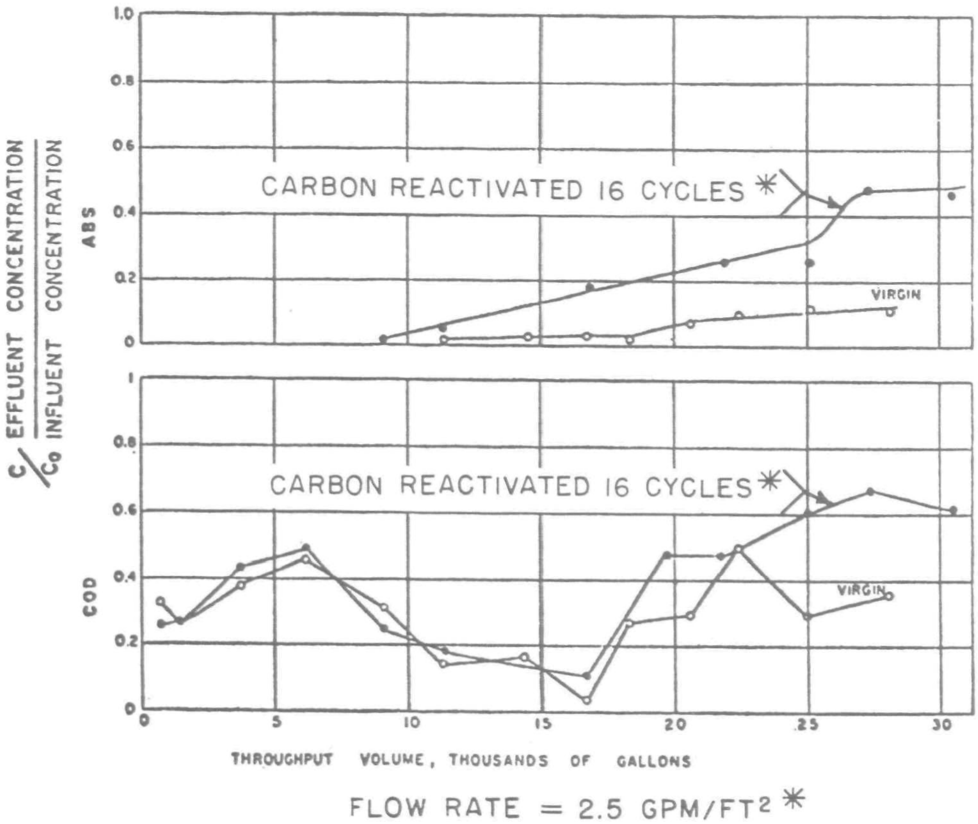
EFFECT OF REGENERATION ON ABS ISOTHERM



REFERENCE 29

FIGURE 27

EFFECT OF REGENERATION ON COD ISOTHERM



REFERENCE 30

FIGURE 28

EFFECT OF REGENERATION ON ABS AND COD REMOVAL EFFICIENCY

REPRINTED BY PERMISSION, Joyce, R.S., *et al.*, *JWPCF*, 38, 818 (1966).

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2. The carbon is not being completely regenerated so that there is still adsorbate left on the carbon after regeneration which, in effect, reduces the capacity of the carbon in subsequent adsorption.

Both degradation and incomplete regeneration have the same effect on efficiency of removal; that is, no effect until the carbon becomes heavily loaded.

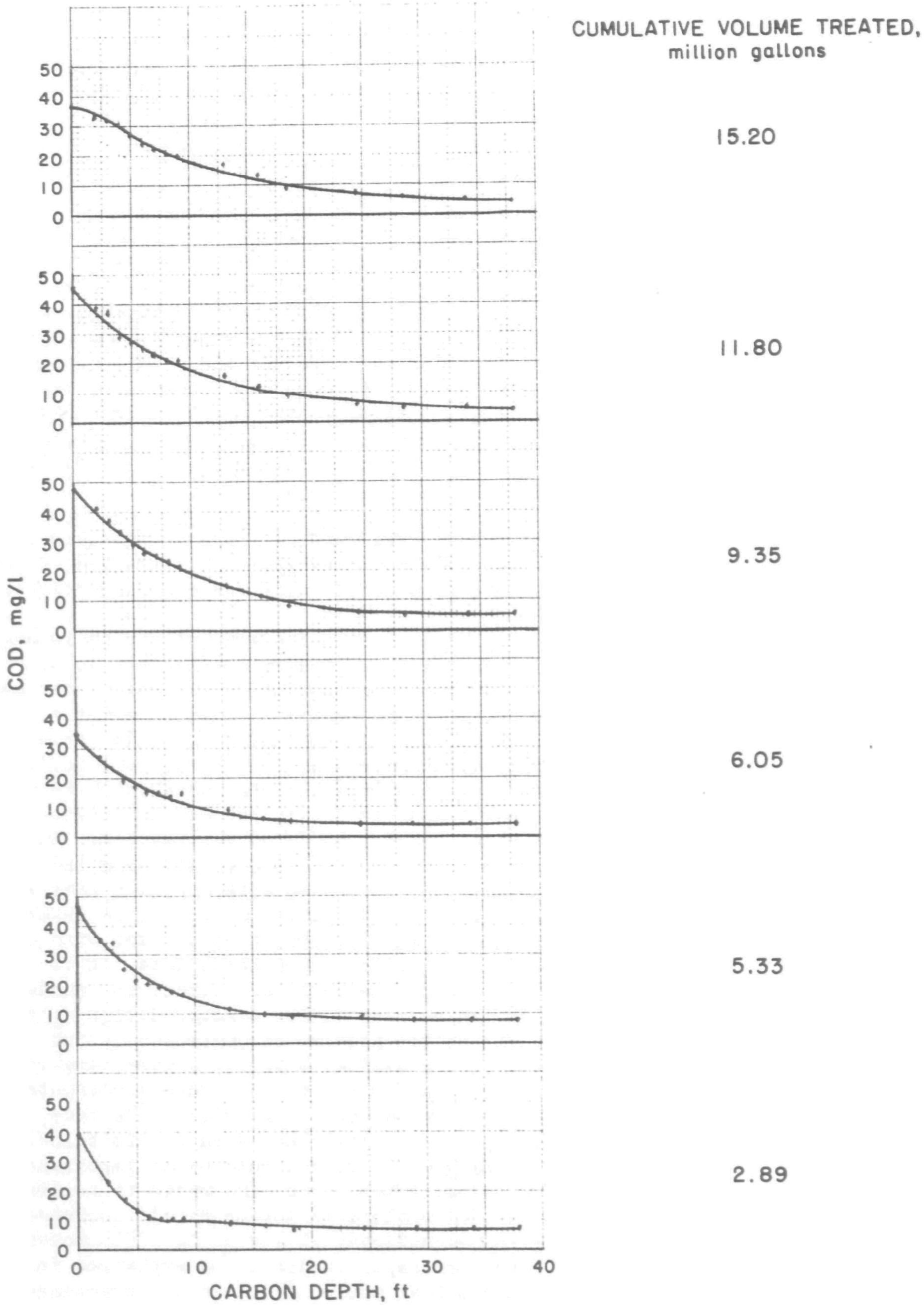
In summary, it appears that the carbon's ability to remove to the desired concentration of COD is not affected by regeneration but that the capacity is decreased as shown in Figure 23. Evidently, the regeneration process affects the equilibrium characteristics of the carbon but does not affect the rate processes.

DISCUSSION

Presented in the previous sections are data relating to the capacity and rate of carbon adsorption in a fixed bed, downflow column and the effects of liquid velocity, particle size and regeneration on the rate and capacity. These data have all been extracted from the literature on tertiary treatment - no new experimental work has been performed. However, a good deal of the data treatment described herein represents independent correlations made in an effort to provide fundamental engineering bases for subsequent process designs. In addition, since the basic data have been obtained by a rather large number of investigators, a considerable amount of effort was expended to bring as much of the data as possible together on consistent bases.

With these data and correlations available, it appears possible to design with a good degree of precision for the case of a fixed bed, downflow column capable of operating over a fairly wide range of conditions.

However, it is felt that the data are not fundamental enough in nature to permit their use as basic design data for system configurations other than the type in which they were taken. This point is in general agreement with others engaged in studies on this subject.[1,31] This is not the fault of the investigators but is a characteristic of the system under study. For example, waste water contains many components all of which, undoubtedly, adsorb at different rates resulting in a breakthrough curve which is not clearly defined (Figure 29). Also, waste water is subject to day-to-day fluctuations in concentration which further confound the results. In addition, capacities greater than the isotherm capacity plus the fact that the carbon could not be completely saturated with COD hint that some mechanism other than adsorption, such as biological activity, is involved.[1] The operation under study is, then, a multicomponent adsorption with complicating factors not related to adsorption. From these data, it is impossible to apply standard adsorption analysis and extract any fundamental mass transfer data which can be used in the precise design of various kinds of contacting systems. The entire process of waste adsorption probably will never lend itself to a classical chemical engineering treatment which will yield fundamental data for any of the process parameters presently under investigation. Before any such analysis can be made, a more basic understanding of the processes involved, such as biological activity, must be developed. Biological action should be an important factor in the removal of waste using carbon, since the waste is concentrated on the carbon and the rate of biological action should increase accordingly. These rates must be significant if the 50 to 100% excess capacity of the carbon above isotherm capacity are to be explained in this way. Also, since the rate of biological action is time- and temperature-dependent in a much different way than is adsorption, this action must be well understood before the entire process can be evaluated. A basic



REFERENCE 32

FIGURE 29

COD BREAKTHROUGH CURVES

understanding of the biological processes in the carbon bed would, hopefully, reveal ways in which the contribution of such processes to the carbon's capacity might be enhanced.

Lacking such basic information, the design engineer is left with only one alternative, that is, using data obtained under controlled conditions from a side-by-side comparison of the different types of contactors. Lake Tahoe has attempted such a comparison of upflow and downflow, fixed-bed contactors[33] but their results were clouded by plugging problems and by running the downflow column at only 2 GPM/ft^2 and the upflow column at 6 GPM/ft^2.

In summary, the data presented herein are sufficient to permit the design of the downflow, fixed-bed contactor but are not sufficient to permit the design of any other type of system with a comparable degree of precision. However, the data are adequate for use in preliminary screening designs of other system configurations (e.g. fludized bed). Such designs may well prove to be useful in determining what potential benefits might be realized by one system over another and would also serve to aid in planning future experimental work. Since there are now no consistent comparative economic (or technical) evaluations for the various proposed schemes - even preliminary ones - such a study would be beneficial.

RECOMMENDATIONS

Using data presently available, preparation of preliminary process designs and evaluations for the various types of proposed contacting systems should begin. Such studies will hopefully determine which scheme has the maximum technical and economic potential as well as indicating what experimental data - if any - are required to bring it to commercial reality.

However, in view of the deficiencies in the present data, there are several gaps which must be filled before a precise economic optimization and comparison can be made of the several contacting systems.

1. A side-by-side experimental comparison of upflow and downflow columns should be made in extended runs with regeneration. The scale of the test should be large enough to determine the effect, if any, of backmixing of the carbon during adsorption and removal in the partially-expanded, upflow bed. A study of the effect of contact time (bed depth) on effluent purity should show the difference in efficiency between these columns. Studies of carbon losses by attrition must also be made.

2. The effect of velocity below 4 GPM/ft^2 must be determined if gravity flow systems are to evaluated. Also, the effect of high velocities on removal efficiency of COD must be obtained, in order to determine if the advantage of high velocity shown in Figure 12 still exists at long contact times. If this is true, reductions in investment could be realized by running at high velocities.

3. The effect of particle size on adsorption capacity and rate at long contact times will determine if the trend shown in Figure 16 will continue or if the difference between large and small carbons will diminish as is expected (Figure 14). The effect of particle size after regeneration should also be studied.

4. The effect of number of regeneration cycles on capacity and rate of adsorption will show if the trend shown in Figure 23 will eventually level off as expected. These data will be

useful in optimization of the plant with regard to carbon investment and length of its use. The effect of velocity after regeneration should also be studied.

5. The effect of particle size on regeneration loss should be determined in extended continuous runs under the same conditions of regeneration. This effect has a direct effect on optimization relative to particle size.

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PHASE II. ECONOMIC EFFECT OF DESIGN VARIABLES

by

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SUMMARY

This report summarizes the findings of work on tertiary waste water treatment using granular activated carbon. Specifically this study found the economic importance of several design variables and showed where additional experimental data are needed. The variables which were studied for their effect on the economics were shop fabrication and field erection of vessels, surge designs, plant size, idle carbon inventory, velocity, contact time, particle size, regeneration loss, carbon capacity, downflow, upflow, gravity-flow and moving-bed contactors, adsorbent cost, number of contacting stages, in-place regeneration and certain combinations of the above variables. The comparisons made in this report are based on data which were sometimes rather fragmentary as found in Phase I of this work.[1] In this respect, the comparisons had to be limited by the inability to get complete predictions of the effect of the variables on the process performance. The assumptions which were made and the recommendations for further work are listed in the body of the report.

Based on comparison of investment and operating costs, several conclusions were drawn. Among the more important conclusions are the following:

1. The gravity-flow system is less expensive than the pressurized downflow system by two cents per thousand gallons which, in turn, is less expensive than the pressurized upflow system with pre-filtration by one cent per thousand gallons. Assumptions made in the case of the gravity-flow contactor are that the required removal will be achieved at 2 GPM/ft^2 with 50-minute contact time and that the loading achieved in a single-stage system is the same as that used in the design of the two-stage system.

2. A two-stage contactor is an optimum system as compared with single-, three-, and four-stage contactors. Large-scale operating data are required to find the allowable carbon adsorption capacity in a two-stage system.

3. Adsorbent cost and regeneration loss have a significant effect on total operating cost. Operating costs would be reduced by 1.15 cents per thousand gallons if the adsorbent costs could be cut by 15 cents per pound. If the regeneration loss is reduced from 10% to 2%, the operating costs would be reduced by two cents per thousand gallons. Efforts should be made to find a lower cost adsorbent, to reduce carbon inventory, to pin-point where regeneration loss occurs and to understand the regeneration mechanism.

4. For a pressurized vessel system, large particle size and high velocity offer savings in costs. Savings in operating cost are possible when 8 x 30 mesh carbon is used rather than 12 x 40 mesh carbon. Also, operating costs can be reduced by 1.7 cents per thousand gallons when the liquid velocity is increased from 4 GPM/ft^2 to 7 GPM/ft^2 Data are required to confirm the effect of velocity on required contact time. Also, it is necessary to find the relationship between particle size and regeneration loss.

5. A large economic incentive exists to try to maximize the carbon loading (capacity). Operating costs can be cut by two cents per thousand gallons when the carbon loading is increased from 0.25 to 0.87 pounds Chemical Oxygen Demand per pound of carbon. This might include enhancement of the biological action by such means as injecting air (oxygen) into the waste water.

INTRODUCTION

The purpose of this report is to summarize the findings of the second phase of work on tertiary waste water treatment using granular activated carbon. The first phase was a review of the literature with a view towards generating design data.[1] The work reported herein represents the results of preliminary economic evaluations of several contacting schemes with a view toward finding which variables have economic significance and showing where additional experimental data are required.

Design variables which were studied for their effect on the economics are listed below.

1. Shop fabrication and field erection of vessels
2. Surge design
3. Plant size
4. Idle carbon inventory
5. Velocity
6. Contact time
7. Particle size
8. Regeneration loss
9. Carbon capacity
10. Downflow, upflow, gravity-flow and moving-bed contactors
11. Adsorbent cost
12. Number of contacting stages
13. In-place regeneration

Table 1 lists all of the cases and design conditions considered. In the absence of correlations between design variables and process performance, various assumptions are made for these correlations to determine the effect on the economics. In this way, each variable is changed one at a time and, if it is found to be economically significant, a recommendation is made to confirm the assumption. A summary table of investment and operating costs has purposely <u>not</u> been provided in order to preclude the making of invalid comparisons. Economic comparisons between cases considered in this study and other cases should not be made unless care is taken to make sure that the cases in question are on

TABLE 1

SUMMARY OF CASES

Case	Contactor Type	Plant Size MGD	Velocity GPM/ft^2	Contact Time Min.	Particle Size Mesh	Regen. Loss %	Carbon Cap. lb COD/lb C	Variable Tested
1	Down	10	7	50	8 x 30	5	0.522	Base Case-Shop Fabrication
1A	Down	10	7	50	8 x 30	5	0.522	Base Case-Field Erection
2	Down	1	7	50	8 x 30	5	0.522	Plant Size
3	Down	100	7	50	8 x 30	5	0.522	Plant Size
3A	Gravity	100	2	50	8 x 30	5	0.522	Contactor Type
4	Down	10	10	50	8 x 30	5	0.522	Δ P & L/D
4A	Down	10	10	50	8 x 30	5	0.522	Idle Carbon
5	Down	10	10	35	8 x 30	5	0.522	Δ P, L/D & C.T.
5A	Down	10	10	35	8 x 30	2	0.522	Improved Downflow
6	Down	10	4	50	8 x 30	5	0.522	Δ P, & L/D
7	Down	10	4	87	8 x 30	5	0.522	Δ P, L/D & C.T.
8	Down	10	7	50	12 x 40	5	0.522	Δ P & Carbon Cost
9	Down	10	7	50	12 x 40	10	0.522	Δ P & Regen. Loss
10	Down	10	7	40	12 x 40	5	0.522	C.T. & Carbon Cost
10A	Down	10	7	40	12 x 40	5	0.522	Idle Carbon
11	Down	10	7	40	12 x 40	10	0.522	Regen. Loss, C.T., Cost & ΔP
12	Down	10	7	50	8 x 30	2	0.522	Regen. Loss
12A	Down	10	7	50	8 x 30	10	0.522	Regen. Loss
13	Down	10	7	50	8 x 30	5	0.87	Capacity
14	Down	10	7	50	8 x 30	5	0.25	Capacity
15	Up	10	7	50	8 x 30	5	0.348	Contactor Type
16	Gravity	10	2	50	8 x 30	5	0.522	Contactor Type
17	Down	10	7	50	8 x 30	5	0.522	Adsorbent Cost
18	Moving-bed	10	7	15	8 x 30	5	0.25	Lake Tahoe System
19	Moving-bed	10	10	15	8 x 30	2	0.348	Improved Moving-Bed
20	Down	10	7	50	8 x 30	5	-	4-Stage System - Pomona System
20A	Down	10	7	50	8 x 30	5	-	3-Stage System
20B	Down	10	7	50	8 x 30	5	-	1-Stage System
21	Down	10(15)	7	50	8 x 30	5	0.522	Surge Design
22	Down	10	7	50	8 x 30	0	0.522	In-Place Regen.

consistent bases. The comparisons which are valid have been made in this study and any other comparison between cases shown should not be made. The accuracy of the investment and operating costs will be improved by the more definitive process design and cost estimate planned for the third and last phase of the contract. Nevertheless, the current accuracy is judged to be sufficient to show trends and to narrow down the choice of process conditions to be used in the definitive design of the next phase.

The estimated cost for the carbon treatment plants is based on the assumptions that the plants will be designed, procured, and constructed by one general contractor; that the geographical location is the Gulf Coast area for the purpose of establishing environmental conditions, labor costs and freight shipment rates; and that the prices of materials equipment and labor are as of November 1968.

Also, when examining the operating costs, it should be remembered that for a 10 million gallon per day (10-MGD) plant, each one cent per thousand gallons (1 cent/M gal) is equivalent to $36,500/year in operating costs. Approximately $230,000 in plant investment can be spent in order to make a savings in operating costs of 1 cent/M gal. Fixed charges are 15.6% per year of the plant investment plus 8.4% per year of the carbon investment.

It should also be noted that an optimum carbon adsorption process does not necessarily mean an optimum total waste treatment process. There may be combinations or interactions between sections of the total waste treatment process which would serve to lower the entire treatment costs. Interactions might be found between the secondary and carbon treatment plants where the operation of the secondary plant would be altered to improve the performance of the carbon plant, thereby reducing the total treatment cost.

References in this report to figures, tables and conclusions in the literature report refer to the Phase I Report which precedes this.

GENERAL PROCESS DESCRIPTION

Following is a process description of the base case (Case 1A) flow sheet (Figure 1) which is typical for a 10 million gallon per day (MGD) plant. The flow sheet for individual cases will not vary substantially from this case. These variations will be discussed with each case.

Effluent from the secondary treatment plant (activated sludge) enters the tertiary treatment plant into a 1.6 million gallon surge basin and at a flow rate varying between 3,470 and 10,410 GPM. Suction for the feed pump, J-1, is taken from the surge basin at a rate of 6940 GPM. This feed water contains 60 mg/L total Chemical Oxygen Demand (COD), 70% of which is dissolved COD and 30% suspended COD. Water from the feed pump then flows through two adsorbers, D-1a and b, in series where 50-minute residence time of the water in the packed carbon beds is provided to produce a product water with 9.5 mg/L total COD and 7 mg/L dissolved COD. This remaining COD is principally dissolved organic material which cannot be readily adsorbed on activated carbon.

Periodically, a carbon bed stops producing water of the desired purity and the carbon in the lead contactor must be regenerated. At this time, the contactor (D-1e on the flow sheet) is taken offstream, a contactor in the number two position is switched to the lead position, a contactor containing regenerated carbon is brought onstream in the number two position and the carbon is transported to a regeneration furnace at the rate of 350 pounds per hour on a continuous basis. To accomplish this transfer the carbon is educted with water through spent carbon eductor, L-2e, from the bottom of the adsorber to the furnace area. The carbon is separated from the water in a dewatering screw conveyor, V-1, and is sent to the regeneration system, L-1, where the carbon is fed to the top of a multiple-hearth, Herreshoff-type regeneration furnace where hot flue gas and steam are used to regenerate the carbon. The overflow water from the screw conveyor is returned to the settlers in the primary treatment plant. The vent gas from the furnace is sent to a cyclone, L-3, and afterburner, L-4, for air pollution control. The hot, regenerated carbon is quenched with water in the quench tank and educted as a slurry back to an adsorber, D-1f, which has been partially filled with water to prevent attrition of the carbon as it falls into the adsorber. The excess water and carbon fines overflow from the top of the adsorber and are sent to the primary settlers. Since carbon is lost in the regeneration process, make-up carbon from the fresh carbon storage bin, F-1, is added to the carbon-water slurry in the quench tank at the rate of 420 pounds per day.

One additional operation which must be performed is that of backwashing the lead adsorbers. Whenever the pressure drop in the system becomes high due to the build-up of suspended solids on the top layer of the carbon, the adsorber (D-1c on the flow sheet) is taken offstream and backwashed with secondary effluent from back wash pump, J-2. The backwash flow rate is set at 17 GPM/ft^2 so that there is a 30% bed expansion during backwashing.

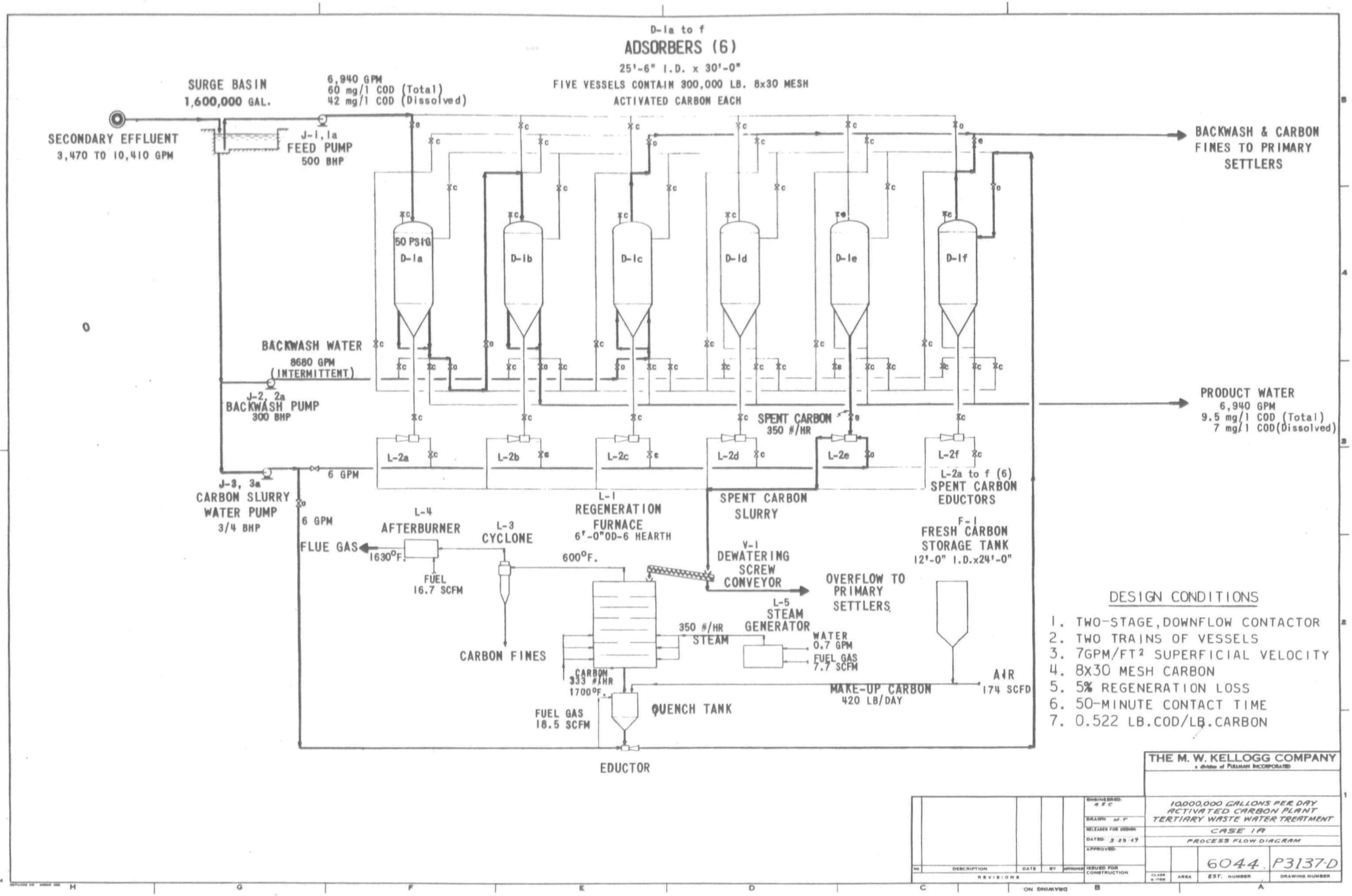
D-1a to f
ADSORBERS (6)
25'-6" I.D. x 30'-0"
FIVE VESSELS CONTAIN 300,000 LB. 8x30 MESH
ACTIVATED CARBON EACH
SURGE BASIN
1,600,000 GAL.
6,940 GPM
60 mg/l COD (Total)
42 mg/l COD (Dissolved)
SECONDARY EFFLUENT
3,470 TO 10,410 GPM
J-1, 1a
FEED PUMP
500 BHP
BACKWASH & CARBON FINES TO PRIMARY SETTLERS
50 PSIG
D-1a
D-1b
D-1c
D-1d
D-1e
D-1f
BACKWASH WATER
8680 GPM
(INTERMITTENT)
J-2, 2a
BACKWASH PUMP
300 BHP
PRODUCT WATER
6,940 GPM
9.5 mg/l COD (Total)
7 mg/l COD (Dissolved)
SPENT CARBON
350 #/HR
L-2a
L-2b
L-2c
L-2d
L-2e
L-2f
L-2a to f (6)
SPENT CARBON EDUCTORS
6 GPM
J-3, 3a
CARBON SLURRY WATER PUMP
3/4 BHP
6 GPM
L-4
AFTERBURNER
L-3
CYCLONE
L-1
REGENERATION FURNACE
6'-0"OD-6 HEARTH
SPENT CARBON SLURRY
F-1
FRESH CARBON STORAGE TANK
12'-0" I.D.x24'-0"
FLUE GAS
1630°F.
FUEL
16.7 SCFM
600°F.
V-1
DEWATERING SCREW CONVEYOR
OVERFLOW TO PRIMARY SETTLERS
L-5
STEAM GENERATOR
350 #/HR
STEAM
WATER
0.7 GPM
FUEL GAS
7.7 SCFM
CARBON FINES
CARBON
333 #/HR
1700°F.
MAKE-UP CARBON
420 LB/DAY
AIR
174 SCFD
FUEL GAS
18.5 SCFM
QUENCH TANK
EDUCTOR
DESIGN CONDITIONS
1. TWO-STAGE, DOWNFLOW CONTACTOR
2. TWO TRAINS OF VESSELS
3. 7GPM/FT² SUPERFICIAL VELOCITY
4. 8X30 MESH CARBON
5. 5% REGENERATION LOSS
6. 50-MINUTE CONTACT TIME
7. 0.522 LB.COD/LB.CARBON
THE M. W. KELLOGG COMPANY
a division of PULLMAN INCORPORATED
10,000,000 GALLONS PER DAY
ACTIVATED CARBON PLANT
TERTIARY WASTE WATER TREATMENT
CASE 1A
PROCESS FLOW DIAGRAM
6044
P3137-D
EST. NUMBER
DRAWING NUMBER
REVISIONS

FIGURE 1

DESIGN BASES

The following items are the design bases used for the base case (Cases 1 and 1A). The bases vary somewhat from case to case in each comparison; the difference in bases from the base case will be discussed.

Water Quality - The Chemical Oxygen Demand (COD) concentration to the activated carbon adsorption process is assumed to be 60 mg/L with other water properties the same as that found at Pomona, California. Suspended solids comprise about 30% of the total COD with the remainder being dissolved COD.

Plant Size - A plant capable of treating 10 million gallons per day (10-MGD) of waste water was chosen because 10-MGD represents the sewage rate from cities of about 100,000 to 300,000 population.

Surge Capacity - A surge basin will be provided to damp-out variations in feed rate. The flow rate is assumed to vary sinusoidally between 50% and 150% of the average flow rate with a period of one day.

Vessel System - A two-stage, packed-bed, downflow system was selected for the base case. The packed-bed, downflow adsorber was chosen because most of the pilot plant data were taken on this type of contactor. A two-stage contactor rather than one-, three-, or four-stages was chosen because it was felt that his system should yield a lower investment. Reductions are expected in vessel and piping costs which should offset operating costs associated with a lower carbon loading for the two-stage system as compared to the four-stage system. Depths of pressure vessels are 30% greater than the carbon bed depths to provide for bed expansion during backwashing.

Carbon Particle Size - The 8 x 30 mesh carbon rather than the 16 x 40 mesh used at Pomona was chosen for the base case because this larger size may reduce the carbon loss during regeneration while not greatly reducing the capacity. Carbon physical characteristics are assumed to be the same as the Pittsburgh Activated Carbon Company's type SGL (8 x 30 mesh) and type CAL (12 x 40 mesh).

Velocity - The superficial linear velocity (based on empty column area) of the water moving through the carbon bed is taken at 7 GPM/ft^2. This velocity was chosen for the base case because most of the pilot plant data were taken at this velocity.

Contact Time - The required contact time to reduce the COD from 60 to 7 mg/L is set at 50 minutes (based on empty column volume). This is longer than the 41-minute contact time at Pomona but a longer contact time is consistent with the use of a larger particle size[2]. As more data become available, optimized design may permit within certain limitations, a reduction in contact time coupled with a higher regeneration rate.

Carbon Bed Dimensions - A minimum bed depth-to-diameter ratio (L/D) on the order of one in a vessel was chosen to assure good contacting during adsorption and backwash. A minimum ratio of this order of magnitude has been found to be required in order to prevent backmixing of the carbon during backwash and to prevent channeling during adsorption. Channeling effects are a function of velocity and would not be as great a problem at the low velocities encountered in gravity-flow systems as compared to pressurized, high- velocity contactors. That is, less pressure drop would be required to overcome channeling problem at low flow rates than at high flow rates. The contactors at Pomona and Lake Tahoe have bed depth-to-diameter ratios greater than one with velocities at about 7 GPM/ft^2 while the contactors at Nitro have a ratio less than one with a velocity at about 1 GPM/ft^2. Pittsburgh Activated Carbon Company in their "Basic Design Techniques" bulletin recommend an L/D ratio of two, but in their report to USPHS[3], they used ratios less than one. An L/D ratio of one should be adequate to assure good contacting for packed columns operating under flooded conditions at high velocities.

"Idle" Carbon Inventory - The portion of the total carbon inventory which must be in residence in the system for the purpose of carrying out regeneration without upsetting the normal operation of the adsorption train will be referred to as "idle" carbon. This carbon is idle with respect to the adsorption of waste matter but, on the other hand, is not idle in the regeneration system. The amount of "idle" carbon required for regeneration purposes is equal to the amount in one vessel. The amount of carbon held-up in the regeneration furnace is less than 1% of the total carbon inventory and has been neglected.

Backwash Rate - The backwash flow rate was chosen to assure a 30% bed expansion during backwash. This is equivalent to the bed expansion employed at Pomona. Good backwashing should be a function of a degree of bed expansion rather than of water velocity so that the accumulated suspended solids can disengage from the interstices of the carbon particles. As at Pomona, backwashing of the lead contactor is assumed to occur once a day using 5% of the daily output from the secondary plant for backwash and surface wash. Bed expansion during backwashing is taken from Cooper and Hager.[4]

Pressure Drop - For the purpose of feed pump sizing, a 50-psi pressure drop is assumed to be developed before backwashing is required. This is approximately the same as at Pomona even though the particle size at Pomona is smaller (16 x 40 vs. 8 x 30). The major portion of the bed pressure drop is contributed by the suspended solids accumulated on the top layer of carbon. This pressure drop will not change much over this particle size range since the void fraction for 12 x 40 is 0.38 and for 8 x 30 is 0.36, not a significant difference. Carbon bed pressure drop is taken from Cooper and Hager.[4] A spare pump is included for each pump. A gas engine or diesel drive is provided for the spare feed and backwash pump so that the plant would not have to be shut down in the case of power failure.

"Spare Vessels"- Two extra vessels are added to the number of vessels which are onstream in the adsorption train. The two "spare" vessels are required for the operation of the regeneration system to store the "idle" carbon which is ready to be and has been regenerated. These two vessels which are in regeneration service are only "spare" insofar as they are not onstream in the adsorption train.

Regeneration Rate - Regeneration of the spent carbon is assumed to occur continuously in a multiple-hearth, Herreschoff-type furnace. The regeneration rate is set by the material balance around the plant and an assumed carbon loading of 0.522 lb COD/lb carbon. This loading corresponds to the loading at Pomona after the carbon has been regenerated ten times[5]. It is assumed that the loading for the two-stage system will be the same as the loading obtained in the four-stage system at Pomona. This assumption will be tested for its effect on the economics in a later section. The regeneration of carbon from the vessels is assumed to occur on a staggered basis so that all of the lead adsorbers do not have to be regenerated at the same time. Staggering of regeneration in a multi-trained system presents the possibility of blending product waters which are worse than and better than specified quality. This allows the carbon to get more heavily loaded than in a single-train system thus reducing the regeneration rate.

Carbon Transport - The spent carbon is transported between the vessel and the furnace area as a carbon-water slurry. The slurry is then drained in an inclined dewatering screw coveyor (as at Nitro, West Virginia) where a 10-minute dewatering period is provided. Dewatering studies at Lake Tahoe indicate that a moisture content of 45% can be achieved after a 10-minute drain time while only 40% moisture can be reached after draining for one day[6]. Because of the two "spare" vessels and the feasibility of carbon dewatering in an inclined conveyor, there is no need for elevated dewatering bins as are in use at Pomona and Lake Tahoe.

The carbon slurry transport system design was based on recommendations in the Pittsburgh Activated Carbon Company's Bulletin, "Column Operating Procedures with Pittsburgh Granular Activated Carbon". Their recommendations are:

1. Minimum linear velocity of 3 feet per second in slurry lines to prevent carbon settling.

2. A carbon-to-water ratio of one pound carbon per gallon of water.

3. Pipeline pressure drop shown in Figure 4 in the above bulletin.

Transportation of the make-up carbon from the storage tank to the quench tank is done pneumatically with the design based on recommendations in McCabe and Smith[7]:

1. The air velocity should be between 50 and 100 feet per second.

2. The solids should occupy 3 to 12% of the volume or the solids-to-air mass ratio should be between 30 and 100.

Regeneration Furnace - Continuous regeneration of the spent carbon occurs in a multiple-hearth, Herreschoff-type furnace. The furnace is sized on the recommendation from B-S-P Corporation of 100 pounds carbon regenerated per day per square foot of hearth area. This design point was used along with the hearth areas given in B-S-P Bulletin No. 250. Air pollution control equipment is provided to prevent odors and particulates from escaping in the flue gas stream.

Regeneration Fuel - The total fuel requirement for the regeneration system is estimated to be 6950 Btu per pound of carbon regenerated. This figure is based on the following recommendations:

1. B-S-P recommends 3000 Btu/lb carbon for furnace heat which is confirmed by Pomona operating data.

2. Lake Tahoe data indicate that 1250 Btu/lb carbon for steam generation of one pound steam per pound carbon is sufficient for satisfactory carbon regeneration.

3. The afterburner fuel requirement for odor control was taken at 2700 Btu/lb carbon to raise the flue gas temperature to 1600°F based on Pomona operating data.

Carbon Loss - The carbon loss in regeneration is assumed to be 5% which is the same as that at Tahoe with 8 x 30 mesh carbon. Storage facilities are provided for a six-month supply of make-up carbon which is the amount of carbon in one adsorber in base case 1A.

Materials of Construction - The materials of construction specified for the purpose of comparative cost estimation is coal tar-epoxy lined carbon steel vessels with carbon steel piping, the materials which are in use at Pomona. In light of recent corrosion information, however, a sheet lining such as laminated hard rubber or PVC would be recommended. The failure of the coal tar-epoxy linings at Pomona and Tahoe is believed to be due to poor application, resulting in pin-holes. The use of new materials would not affect the economic comparisons made in this report, but would raise the capital cost of all vessels.

Operating Labor Requirement - A labor requirement of 1-1/4 man per shift for operation of a 10-MGD plant is estimated -- one man operating the regeneration system and one man monitoring vessel operation one-quarter of the time.

Calculation of Other Operating Costs - The basis and rationale used in the procedure for calculation of other operating costs such as maintenance, overhead, amortization, insurance, taxes and unit costs for utilities are given in Appendix A.

Plant Investment - The items which are included in the estimation of the plant investment are given in Appendix A.

ECONOMICS

Comparison of Shop Fabrication and Field Erection of Vessels
Cases 1 & 1A

The comparison of investment and operating costs for the shop - fabricated (Case 1) and field-erected vessels (Case 1A) is shown below. The limitation of 13 feet in vessel diameter in the shop fabrication case resulted in 18 vessels required for the 10-MGD plant (8 trains of 2 vessels each plus 2 spare vessels.) The field erection of vessels coupled with the restriction of a minimum bed height-to-diameter ratio of one resulted in six 25'6" I.D. vessels. In the field erection case, a larger backwash pump is required in order to provide the 30% bed expansion for good backwashing. Cost comparisons are presented in Table 2 for shop fabrication and field erection. The table shows no significant difference in either investment or operating costs between the two methods of construction. Savings in field erection of vessel steel and piping are offset by field labor, backwash pump and carbon costs. The process simplicity of field erection made it a logical choice for the basis of subsequent comparisons.

Comparison of Surge Design and Base Case
Cases 1A & 21

The question being tested in this comparison is whether it is cheaper to handle the surge in flow rate from 5 to 15 MGD by damping the variation in a surge basin (Case 1A) or by bringing extra vessels onstream (Case 21). It was assumed that this variation in flow rate would occur sinusoidally with a period of one day. A 1,600,000-gallon basin is required to store the excess water while in the other case, three trains of vessels are required (two additional vessels). The costs for this comparison are presented in Table 3 which show that it is significantly more expensive to handle surges in flow by adding vessels. The two extra vessels along with the extra carbon, piping and larger feed pump cost more than does the concrete surge basin. The higher operating costs are a result of the higher investment. The utility costs average out to be the same as the base case.

Effect of Plant Size on Economics
Cases 1A, 2 & 3

These comparisons attempt to find what economies of scale, if any, may be realized. The 1-MGD plant requires four 11'6" I.D. vessels, the 10-MGD plant, six 25'6" I.D. vessels and the 100-MGD plant, forty-four 25'6" I.D. vessels (20 trains of two vessels each plus four spare vessels). The regeneration furnace for the 1-MGD plant is the minimum

TABLE 2

COMPARISON OF SHOP FABRICATION AND FIELD ERECTION OF VESSELS

Case No.	1	1A
Variable Tested	Shop Fabrication	Field Erection
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	50
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	13' I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	18	6
Number of Trains	8	2
Investment, $M		
Concrete	126	121
Adsorbers	385	253
Tanks	5	5
Pumps	68	98
Special Equipment	40	36
Piping	213	158
Conveyors	4	4
Total Major Material	841	675
Plant Investment	1294	1210
Carbon @ 26¢/lb.	345	390
Fixed-Capital Investment	1639	1600
Operating Costs, ¢/M Gal.		
Make-up Carbon @ 26¢/lb.	1.09	1.09
Power @ 1¢/kwh	1.01	1.00
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.05	1.02
Amortization @ 7.4% FCI/Yr.	3.32	3.24
Maintenance @ 5.75% PI/Yr.	2.04	1.91
Insurance @ 1% FCI/Yr.	0.45	0.44
Total Operating Cost	10.47	10.21

TABLE 3

COMPARISON OF SURGE DESIGN AND BASE CASE

Case No.	1A	21
Variable Tested	Surge Basin	Extra Vessels
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	50
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	25'6" I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	8
Number of Trains	2	3
Investment, $M		
Concrete	121	60
Adsorbers	253	337
Tanks	5	5
Pumps	98	129
Special Equipment	36	37
Piping	158	210
Conveyors	4	4
Total Major Material	675	782
Plant Investment	1210	1431
Carbon @ 26¢/lb.	390	546
Fixed-Capital Investment	1600	1977
Operating Costs, ¢/M Gal.		
Make-up Carbon @ 26¢/lb.	1.09	1.09
Power @ 1¢/kwh	1.00	1.00
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.11
Amortization @ 7.4% FCI/Yr.	3.24	4.01
Maintenance @ 5.75% PI/Yr.	1.91	2.25
Insurance @ 1% FCI/Yr.	0.44	0.54
Total Operating Cost	10.21	11.51

size (30-inch I.D. - 6 hearth) and would be run only about 35% of the time. The fuel required for start-up and shut-down of the furnace is negligible (4% of the total fuel required) compared to the total consumption. Since the regeneration furnace is run only part of the time, it was assumed that only one man per shift would be required for the entire plant.

The regeneration furnace for the 100-MGD plant is a 16' O.D. -6 hearth furnace with a wet scrubber for air pollution control. B-S-P recommends the use of a wet scrubber rather than a cyclone and afterburner for air pollution control for large furnaces such as this. The operating labor requirement for this plant is three men per shift.

The cost comparison for plant sizes are presented in Table 4, which shows that the investment charges and operating labor are substantially lower in the 10-MGD case than in the 1-MGD case. High labor and investment charges are typical for small plants. On the other hand, the reduction in investment charges from the 10-MGD plant to the 100-MGD plant is not very great because there are multiple trains of equipment required in the 100-MGD case. Accordingly, savings in investment charges are not great when multiple trains are required. The power requirements for the three plant sizes are not the same, principally because the smaller pumps for the smaller plants are less efficient than in the large plants. Also, the power requirement for the regeneration system does not increase in direct proportion to the plant size.

The investment and operating costs are plotted as a function of plant size in Figure 2. A break in the lines is seen at a plant size of 10-MGD because at plant sizes larger than this, the vessels would be erected in the field.

Effect of Gravity-Flow Contactor At Two Plant Sizes on Economics Cases 1A, 16, 3 & 3A

This comparison, which identifies savings which can be expected by replacing a steel pressure vessel contactor system with a concrete gravity-flow contactor system, is made at flow rates of 10 and 100-MGD. A major assumption in the design of the gravity-flow system is that the same contact time is required at 2 and 7 GPM/ft.^{2} to accomplish the same removal of COD. There are data to support this assumption - at least in the range of 4 to 10 GPM/ft^2.[8] Designing a single-stage, gravity-flow contactor for a higher velocity than 2 GPM/ft.^{2} would result in deeper carbon beds and concrete walls which would be prohibitively thick. The contactor walls must be able to withstand the full water pressure since the adjacent contactor will be emptied for regeneration at times. Another assumption in the absence of data is that the same loading will be achieved in the single-stage, gravity-flow system as in the two-stage system. An alternate scheme for the gravity-flow case would be a two-stage system operating at 4 GPM/ft.^{2}. This would avoid the problem with the effect of velocity on rate of adsorption and the effect of a single-stage contactor on loading. The number of contactors would be the same in the single and two-stage system but the height above the carbon for water head would be doubled and two extra headers would be required (one for the outlet

TABLE 4

EFFECT OF PLANT SIZE ON ECONOMICS

Case No.	2	1A	3
Variable Tested	1 MGD	10 MGD	100 MGD
Contactor Type	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7
Contact Time, Min.	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522
Vessel Size	11'6" I.D. x 30'	25'6" I.D. x 30'	25'6" I.D. x 30'
Number of Vessels	4	6	44
Number of Trains	1	2	20
Investment, $M			
Concrete	29	121	811
Adsorbers	74	253	1765
Tanks	2	5	12
Pumps	17	98	433
Special Equipment	27	36	150
Piping	43	158	1599
Conveyors	2	4	23
Total Major Material	194	675	4793
Plant Investment	331	1210	8060
Carbon @ 26¢/lb.	48	390	3276
Fixed-Capital Investment	379	1600	11,736
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	1.09	1.09	1.09
Power @ 1¢/kwh	1.21	1.00	0.79
Backwash Water	0.47(1)	0.31(2)	0.18(3)
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.09
Labor @ $3.50/Man-Hr.	8.40	1.05	0.25
Overhead @ 50% Labor + 1.485% PI/Yr.	5.55	1.02	0.45
Amortization @ 7.4% FCI/Yr.	7.68	3.24	2.38
Maintenance @ 5.75% PI/Yr.	5.21	1.91	1.27
Insurance @ 1% FCI/Yr.	1.04	0.44	0.32
Total Operating Cost	30.80	10.21	6.82

(1) Backwash water (secondary effluent) @ 9¢/M gal.
(2) Backwash water (secondary effluent) @ 6¢/M gal.
(3) Backwash water (secondary effluent) @ 3¢/M gal.

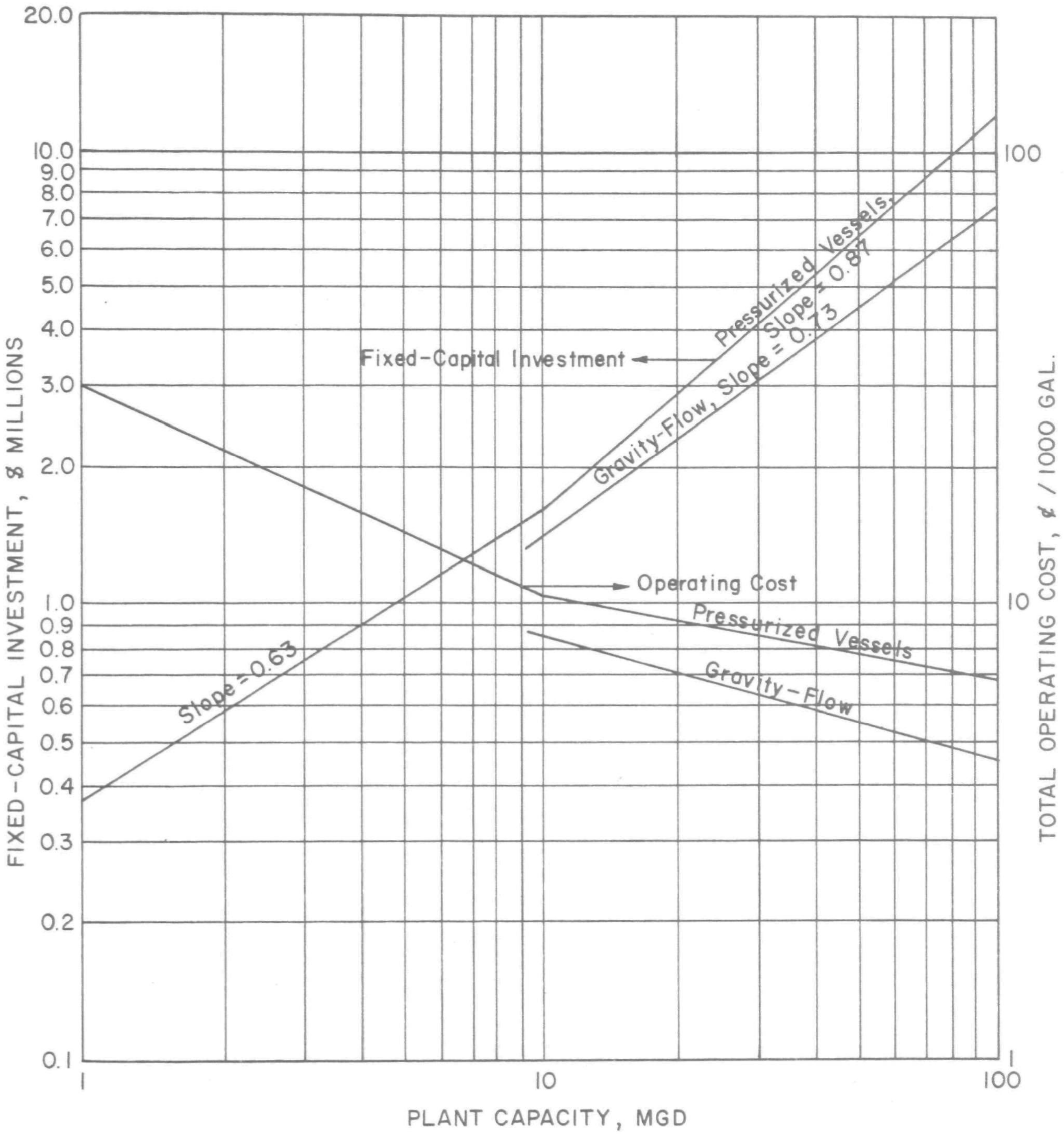

FIGURE 2

COSTS AS A FUNCTION OF PLANT SIZE
TERTIARY WASTE WATER TREATMENT-GRANULAR ACTIVATED CARBON

from the first stage at the bottom of the contactors and another for distribution of this water to the second stage at the grade level). A roof has been included in the cost estimate of the gravity-flow cases for the prevention of algae growth on top of the carbon beds. The requirement of a minimum bed height-to-diameter ratio of about one has been waived in the design of the gravity-flow contactors. For discussion on the bed height-to-diameter ratio, see the "Design Bases" section.

Case 1A is the base case - a pressure vessel system at 10-MGD. The vessel requirement is six 25'6" I.D. x 30' vessels designed for a 95 psi pressure rating. Water discharges from each vessel through eight 8" diameter 40 mesh stainless steel well point type screens.

Case 16 (gravity flow system at 10-MGD) requires 10 rectangular cross section vessels (eight onstream plus two spares), each 20 ft. wide x 22 ft. long x 25 ft.-6 in. high. Each vessel has a 13 ft.-6 in. deep carbon bed with 2 ft. of sand and gravel under the carbon and porous filter bottom under the gravel. A freeboard space of 10 ft. is provided above the carbon bed to allow for bed expansion during backwash and to provide height for the water to rise as the bed pressure drop increases due to the build-up of suspended solids on the bed. Two separate sets of troughs are proveded: one for removal of the backwash water and the other for carbon removal for regeneration. Pumps are provided for pumping the feed water from the bottom of the surge basin to the feed conduit and from the filtered water conduit at the bottom of the contactors back to ground level. If the system had complete gravity flow (e.g., if it were built on a hill), one or both of these pumps would not be needed and would result in savings of investment and power costs.

Case 3 (pressure vessel contacting system at 100-MGD) requires forty-four 25'6" I.D. x 30' vessels (40 onstream plus 4 spares).

Case 3A (gravity-flow at 100-MGD) requires 20 rectangular cross section vessels (18 onstream plus 2 spares), each 40 ft. wide x 50 ft. long x 25 ft.-6 in. high. The contactor construction is the same as in Case 16 described above except that each contactor is split lengthwise by a central bay for water distribution.

The investment and operating costs for these four cases are shown in Table 5; as can be seen, the gravity-flow contactor is significantly less expensive than the vessel system at both 10- and 100-MGD. In Table 5, the concrete cost is for the concrete in place which includes field labor costs. Adsorber cost in Cases 1A and 3 is for material only. It must be re-emphasized that it has been assumed in the absence of data that a 50-minute contact time will accomplish the same COD removal at both 2 and 7 GPM/ft.^{2} and that the 0.522 lb. COD/lb. carbon capacity can be obtained in a single-stage contactor. The savings in the gravity-flow cases are due mainly to savings in investment. Investment savings come about from the fact that concrete contactors are less expensive than steel vessels. The investment and operating costs for gravity-flow and pressurized vessel systems are plotted as a function of plant size in Figure 2.

If the feed water and product pumps could be eliminated for a complete gravity-flow system, then the fixed-capital investment would be reduced by $71,000 and $293,000 for Cases 16 and 3A, respectively. The

TABLE 5

EFFECT OF GRAVITY FLOW CONTACTOR AT TWO PLANT SIZES ON ECONOMICS

Case No.	1A	16	3	3A
	10 MGD	10 MGD	100 MGD	100 MGD
Variable Tested	Vessels	Gravity	Vessels	Gravity
Contactor Type	Downflow	Gravity	Downflow	Gravity
Velocity, GPM/ft^2	7	2	7	2
Contact Time, Min.	50	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522	0.522
Vessel Size	25'6"I.D. x 30'	20'w x 22'1 x 25'6"h	25'6"I.D. x 30'	40'w x 50'1 x 25'6"h
Number of Vessels	6	10	44	20
Number of Trains	2	8	20	18
Investment, $M				
Concrete	121	426	811	1771
Adsorbers	253	-	1765	-
Tanks	5	5	12	12
Pumps	98	116(1)	433	341(1)
Special Equipment	36	37	150	149
Piping	158	95	1599	527
Conveyors	4	4	23	23
Total Major Material	675	683	4793	2823
Plant Investment	1210	1064	8060	4271
Carbon @ 26¢/lb.	390	347	3276	3335
Fixed-Capital Investment	1600	1411	11,736	7606
Operating Costs, ¢/M Gal.				
Make-up Carbon @ 26¢/lb.	1.09	1.09	1.09	1.09
Power @ 1¢/kwh	1.00	0.64(1)	0.79	0.63(1)
Backwash Water	0.31(2)	0.31(2)	0.18(3)	0.18(3)
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.09	0.09
Labor @ $3.50/Man-Hr.	1.05	1.05	0.25	0.25
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	0.96	0.45	0.30
Amortization @ 7.4% FCI/Yr.	3.24	2.24(4)	2.38	1.21(4)
Maintenance @ 5.75% PI/Yr.	1.91	1.68	1.27	0.67
Insurance @ 1% FCI/Yr.	0.44	0.39	0.32	0.21
Total Operating Cost	10.21	8.51	6.82	4.63

(1) Water back to grade level - costs lower for complete gravity flow - see text
(2) Secondary effluent @ 6¢/M gal.
(3) Secondary effluent @ 3¢/M gal.
(4) 30-year plant life assumed - Amortization Rate = 5.8%/yr.

power and investment charges (at 15.6% of investment per year) would decrease by 0.55 and 0.30¢/M gal., respectively, for Case 16 and 0.060 and 0.13¢/M gal., respectively for Case 3A.

Effect of "Idle" Carbon Inventory on Economics Cases 4, 4A, 10 & 10A

The object of this comparison is to determine the optimum amount of idle carbon which can be present in the system. The term "idle" carbon describes the carbon which must be bought with the plant but is not actually onstream. This portion of the total carbon inventory must be in residence in the system for the purpose of carrying out regeneration without upsetting the normal operation of the adsorption train. It has been assumed that to provide smooth operation during the regeneration sequence, one of the two spare vessels in the plant must be filled with carbon. This amount of idle carbon can be decreased at the expense of extra vessels (more trains) and associated piping. As the number of trains is increased, the size of the vessels decreases; therefore, the amount of carbon in a vessel decreases. The vessel and carbon requirements for the four cases considered are shown below:

Case	Vessels	Total Carbon, lb	Idle Carbon, lb	Fraction Idle Carbon
4	4-30'I.D. x 43'6"	1,776,000	592,000	1/3
4A	6-21'I.D. x 43'6"	1,450,000	290,000	1/5
10A	6-25'6"I.D. x 25'	1,111,000	222,200	1/5
10	10-18'I.D. x 25'	996,300	110,700	1/9

The comparison for the effect of idle carbon is shown in Table 6. It should be noted that Cases 4 and 4A are not to be compared with Cases 10 and 10A because of the changes in velocity, contact time and particle size. Valid comparisons can be made only between Cases 4 and 4A and between Cases 10 and 10A where these variables do not change.

As can be seen by comparing Cases 4 and 4A, there is a 10% savings in investment and a 0.6¢/M gal. savings in operating cost when the idle carbon is reduced from 1/3 to 1/5 of the total carbon inventory. However, when Cases 10 and 10A are compared, it is seen that there is little difference in investment or operating costs when the idle carbon is reduced from 1/5 to 1/9 of the total carbon inventory. The conclusion drawn is that the amount of idle carbon must be between 10% and 20% of the total carbon inventory in order to yield minimum costs. The costs can be expected to rise again as more vessels are added and the idle carbon inventory is reduced further. A point will be reached where the decreasing carbon investment will be more than offset by increasing vessel and piping costs.

TABLE 6

EFFECT OF IDLE CARBON ON ECONOMICS

Case No.	4*	4A*	10A*	10*
Variable Tested	1/3 Idle Carbon 1 Train	1/5 Idle Carbon 2 Trains	1/5 Idle Carbon 2 Trains	1/9 Idle Carbon 4 Trains
Contactor Type	Downflow	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	10	10	7	7
Contact Time, Min.	50	50	40	40
Particle Size, Mesh	8 x 30	8 x 30	12 x 40	12 x 40
Regeneration Loss, %	5	5	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522	0.522
Vessel Size	30'I.D. x 43'6"	21'I.D. x 43'6"	25'6"I.D. x 25'	18' I.D. x 25'
Number of Vessels	4	6	6	10
Number of Trains	1	2	2	4
Investment, $M				
Concrete	112	121	121	121
Adsorbers	342	254	246	219
Tanks	5	5	5	5
Pumps	134	101	103	80
Special Equipment	35	36	36	37
Piping	107	158	158	236
Conveyors	4	4	4	4
Total Major Material	739	679	673	703
Plant Investment	1290	1202	1202	1233
Carbon	462 (1)	377 (1)	322 (2)	289 (2)
Fixed-Capital Investment	1752*	1579*	1524*	1522*
Operating Costs, ¢/M Gal.				
Make-up Carbon	1.09 (1)	1.09 (1)	1.22 (2)	1.22 (2)
Power @ 1¢/kwh	1.00	1.00	1.06	1.08
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.05	1.01	1.01	1.03
Amortization @ 7.4% FCI/Yr.	3.55	3.20	3.09	3.09
Maintenance @ 5.75% PI/Yr.	2.03	1.89	1.89	1.94
Insurance @ 1% FCI/Yr.	0.48	0.43	0.42	0.42
Total Operating Cost	10.71*	10.13*	10.20*	10.29*

(1) 8 x 30 mesh make-up carbon @ 26¢/lb.
(2) 12 x 40 mesh make-up carbon @ 29¢/lb.

* Cases 4 and 4A are <u>not</u> to be compared with Cases 10 and 10A. Valid comparisons can be made only between Cases 4 and 4A and between Cases 10 and 10A.

Effect of Velocity on Economics
Cases 1A, 4A & 6

The purpose of this comparison is to find the effect of velocity alone, under the assumption that velocity does not affect the performance (removal efficiency and carbon loading) of the carbon contactor. This assumption is consistent with most of the data on velocity[8]. Three velocities are considered (4, 7 and 10 GPM/ft.^{2}) to find the economic effect of changing the pressure drop and the contactor height-to-diameter ratio.

The vessel requirement for Case 4A (10 GPM/ft.^{2}) is six 21' I.D. x 43'6" vessels, for Case 1A (7 GPM/ft.^{2}) it is six 25'6" I.D. x 30' vessels and for Case 6 (4 GPM/ft.^{2}) it is twenty-eight 13' I.D. x 17'6" vessels. Thirteen trains of vessels were required for Case 6 due to the restriction of a minimum bed height-to-diameter ratio of one. The rationale for this minimum ratio is discussed in the "Design Bases" section.

The carbon bed pressure drop is 5.5 inches H_2O per foot carbon at 10 GPM/ft.^{2}, 4 in./ft. at 7 GPM/ft.^{2} and 2 in./ft. at 4 GPM/ft.^{2}. Carbon bed pressure drop constitutes only a small portion of the total head which must be developed by the feed pump, the remainder being pressure drop due to suspended solids and pipeline.

The comparison to show the effect of velocity on the economics is shown in Table 7. As can be seen from the economics, there is no significant difference between the 10 and 7 GPM/ft.^{2} cases. The power for the pumps is the same in Case 4A as in Case 1A because the larger feed pump in Case 4A was offset by a smaller backwash pump. The backwash pump is smaller because the vessels are smaller in diameter. All other costs are about the same for these two cases.

The 4 GPM/ft.^{2} for Case 6 is substantially higher in cost than for either the 7 or 10 GPM/ft.^{2} case. The major cost items are the vessels and associated piping. Many vessels (28) were required in Case 6 due to the low velocity coupled with the restriction of a minimum bed height-to-diameter ratio of one which is necessary for good contacting and backwashing.

The conclusion drawn from this comparison is that velocity has little effect on the economics, once the velocity is above 7 GPM/ft.^{2}. The major assumption is that velocity does not affect the performance of the contactor. No additional data are needed to verify the effect of velocity alone on the economics.

Effect of Contact Time at High Velocity on Economics
Cases 4A & 5

This comparison identifies economic effect of changing only the contact time from 50 to 35 minutes with the velocity at 10 GPM/ft.^{2}. Such a reduction in required contact time for 87% removal of COD is predicted if the correlation shown in Figure 12 of the literature report[9]

TABLE 7

EFFECT OF VELOCITY ON ECONOMICS

Case No.	4A	1A	6
Variable Tested	10 GPM/ft^2 Velocity	7 GPM/ft^2 Velocity	4 GPM/ft^2 Velocity
Contactor Type	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	10	7	4
Contact Time, Min.	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522
Vessel Size	21' I.D. x 43'6"	25'6" I.D. x 30'	13' I.D.x17'6"
Number of Vessels	6	6	28
Number of Trains	2	2	13
Investment, $M			
Concrete	121	121	141
Adsorbers	254	253	452
Tanks	5	5	5
Pumps	101	98	68
Special Equipment	36	36	43
Piping	158	158	330
Conveyors	4	4	4
Total Major Material	679	675	1043
Plant Investment	1202	1210	1651
Carbon @ 26¢/lb.	377	390	315
Fixed-Capital Investment	1579	1600	1966
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	1.09	1.09	1.09
Power @ 1¢/kwh	1.00	1.00	0.97
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.01	1.02	1.20
Amortization @ 7.4% FCI/Yr.	3.20	3.24	3.99
Maintenance @ 5.75% PI/Yr.	1.89	1.91	2.60
Insurance @ 1% FCI/Yr.	0.43	0.44	0.54
Total Operating Cost	10.13	10.21	11.90

is valid, rather than the correlations in Figures 3-10[8] which showed no effect of velocity on required contact time. The data in Figure 12 were the only data which could be found in a form suitable for analysis on the effect of velocity on efficiency of removal for COD. It should be noted that there was no effect of velocity on Total Organic Carbon removal[10]. The effect of velocity on COD removal still remains to be resolved. This reduction in contact time results in a reduction of vessel size from 21'I.D. x 43'6" to 21'I.D. x 30'.

The economics for these two cases are given in Table 8 which shows that there is a significant reduction in costs when the contact time is reduced. Contactor volume is, of course, directly proportional to contact time so that savings are realized in both vessel cost and carbon inventory. Power consumption is less in the 35-minute contact time case because the pressure loss is less in the shorter bed. This comparison shows that it is important to know the contact time required in order to achieve the desired removal. The assumption in this case was based on the effect of velocity shown in the literature report.[9]

Effect of Contact Time at Low Velocity on Economics
Cases 6 & 7

This comparison finds the economic effect of changing the contact time from 50 to 87 minutes with the velocity at 4 GPM/ft.^{2}. An increase of this order in contact time is required if the correlation in Figure 12 of the literature report[9] is valid rather than the correlations in Figures 3-10[8]. The data in Figure 12 were the only data which could be found in a form suitable for analysis on the effect of velocity on efficiency of removal for COD. Figures 3-10 predicted that there is no effect of velocity on required contact time and is represented here by Case 6. This increase in contact time (Case 7) results in the vessel requirement changing from twenty-eight 13'I.D. x 17'6" vessels for 50-minute contact time to ten 23'6" I.D. x 30' vessels for 87-minute contact time. The number of vessels for the 87-minute contact time could be reduced because the bed depth in a vessel increased, and the design condition of minimum bed depth-to-diameter of one allows the diameter to increase.

The economics for these two cases are given in Table 9 which shows a significant increase in costs when the contact time is increased from 50 to 87 minutes. Again, contactor volume is directly proportional to contact time so that the carbon investment is increased. The plant investment did not increase correspondingly ($1.651 vs. $1.686 MM) because the number of vessels was reduced from 28 to 10. The power consumption is actually less in Case 7 because the area occupied by the vessels is smaller, thus reducing the pipeline pressure drop by more than the increase in pressure drop due to the deeper carbon bed. As in the preceeding section of this report, this comparison shows that it is quite important to determine the relation between contact time and COD removal. The assumption made in Case 7 was based on the effect of velocity shown in Figure 12 of the literature report[9]. Case 6 corresponds to the correlation in Figures 3-10 in the literature report[8] which showed no effect of velocity on contact time.

TABLE 8

EFFECT OF CONTACT TIME AT HIGH VELOCITY ON ECONOMICS

Case No.	4A	5
Variable Tested	50 Minute Contact Time	35 Minute Contact Time
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	10	10
Contact Time, Min.	50	35
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	21' I.D. x 43'6"	21' I.D. x 30'
Number of Vessels	6	6
Number of Trains	2	2
Investment, $M		
Concrete	121	121
Adsorbers	254	210
Tanks	5	5
Pumps	101	92
Special Equipment	36	36
Piping	158	158
Conveyors	4	4
Total Major Material	679	626
Plant Investment	1202	1102
Carbon @ 26¢/lb.	377	265
Fixed-Capital Investment	1579	1367
Operating Costs, ¢/M Gal.		
Make-up Carbon @ 26¢/lb.	1.09	1.09
Power @ 1¢/kwh	1.00	0.97
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @50% Labor+1.485% PI/Yr.	1.01	0.97
Amortization @ 7.4% FCI/Yr.	3.20	2.77
Maintenance @ 5.75% PI/Yr.	1.89	1.74
Insurance @ 1% FCI/Yr.	0.43	0.37
Total Operating Cost	10.13	9.42

TABLE 9

EFFECT OF CONTACT TIME AT LOW VELOCITY ON ECONOMICS

Case No.	6	7
Variable Tested	50 Minute Contact Time	87 Minute Contact Time
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	4	4
Contact Time, Min.	50	87
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	13'I.D. x 17'6"	23'6"I.D. x 30'
Number of Vessels	28	10
Number of Trains	13	4
Investment, $M		
Concrete	141	126
Adsorbers	452	415
Tanks	5	5
Pumps	68	98
Special Equipment	43	37
Piping	330	252
Conveyors	4	4
Total Major Material	1043	937
Plant Investment	1651	1686
Carbon @ 26¢/lb.	315	596
Fixed-Capital Investment	1966	2282
Operating Costs, ¢/M Gal.		
Make-up Carbon @ 26¢/lb.	1.09	1.09
Power @ 1¢/kwh	0.97	0.93
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.20	1.21
Amortization @ 7.4% FCI/Yr.	3.99	4.63
Maintenance @ 5.75% PI/Yr.	2.60	2.66
Insurance @ 1% FCI/Yr.	0.54	0.63
Total Operating Cost	11.90	12.66

Effect of Velocity and Contact Time on Economics
Cases 1A, 5 & 7

With the COD removal held constant at 87% (60 mg/L feed to 7 mg/L effluent), this comparison identifies the full economic impact of varying velocity (4, 7 and 10 GPM/ft.^{2}) with the assumption that velocity and removal efficiency are related as in Figure 12 of the literature report[9]. That correlation predicts that the depth of carbon required for an equal degree of removal is the same regardless of the velocity - at least in the range of 4 to 10 GPM/ft.^{2}. The data presented in the literature report were the only data which could be found in a form suitable for analysis on the effect of velocity on efficiency of removal for COD. This comparison includes the effects of pressure drop and contactor height-to-diameter ratio. Case 5 (10 GPM/ft.^{2} and 35-minute contact time) requires six 21'I.D. x 30' vessels; Case 1A (7 GPM/ft.^{2} and 50-minute contact time) requires six 25'6" I.D. x 30' vessels; Case 7 (4 GPM/ft.^{2} and 87-minute contact time) requires ten 23'6"I.D. x 30' vessels.

The economics for these three cases are given in Table 10. The economics show that the effect of velocity is quite significant if the assumption that velocity affects required contact time is correct. Since contactor volume is directly proportional to contact time, both the vessel cost and carbon inventory decrease with decreasing contact time. This decreasing investment is, of course, reflected in the operating cost items which are related to investment. The power requirement at 10 GPM/ft.^{2} is less than at 7 GPM/ft.^{2} because of a slightly lower pipeline pressure drop and a smaller backwash pump.

This comparison shows quite clearly how important it is to know the relation between velocity and contact time. Based on this comparison, the system should be run at 10 GPM/ft.^{2}. On the other hand, if velocity had no effect on required contact time (see section on "Effect of Velocity") the system should be run between 7 and 10 GPM/ft.^{2}.

Effect of Particle Size on Economics
Cases 1A & 8

This comparison identifies the economic effect of changing only the particle size from 8 x 30 mesh (Case 1A) to 12 x 40 mesh (Case 8). The smaller particle size increases the carbon bed pressure drop to 8.5 in. H_2O/ft. carbon compared to 4 in. H_2O/ft. carbon for 8 x 30 mesh carbon. The contribution of the suspended solids to the total pressure drop is assumed to be the same for both particle sizes, since the void fraction is about the same for both sizes (0.38 for 12 x 40 vs. 0.36 for 8 x 30). The backwash pump is smaller for the smaller size carbon because a lower velocity is required (10 vs. 17 GPM/ft.^{2}) to achieve a 30% bed expansion during backwashing.

TABLE 10

EFFECT OF VELOCITY AND CONTACT TIME ON ECONOMICS

Case No.	5	1A	7
Variable Tested	10 GPM/ft^2 35 Min. C. T.	7 GPM/ft^2 50 Min. C. T.	4 GPM/ft^2 87 Min. C. T.
Contactor Type	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	10	7	4
Contact Time, Min.	35	50	87
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522
Vessel Size	21'I.D. x 30'	25'6"I.D. x 30'	23'6"I.D. x 30"
Number of Vessels	6	6	10
Number of Trains	2	2	4
Investment, $M			
Concrete	121	121	126
Adsorbers	210	253	415
Tanks	5	5	5
Pumps	92	98	98
Special Equipment	36	36	37
Piping	158	158	252
Conveyors	4	4	4
Total Major Material	626	675	937
Plant Investment	1102	1210	1686
Carbon @ 26¢/lb.	265	390	596
Fixed-Capital Investment	1367	1600	2282
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	1.09	1.09	1.09
Power @ 1¢/kwh	0.97	1.00	0.93
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05
Overhead @50% Labor+1.485% PI/Yr.	0.97	1.02	1.21
Amortization @ 7.4% FCI/Yr.	2.77	3.24	4.63
Maintenance @ 5.75% PI/Yr.	1.74	1.91	2.66
Insurance @ 1% FCI/Yr.	0.37	0.44	0.63
Total Operating Cost	9.42	10.21	12.66

The bulk density for the 12 x 40 mesh carbon is less than for 8 x 30 mesh (22.9 vs. 25.0 lb./ft.^{3} thus, a smaller quantity of carbon is required to provide the same contact time, however, this lower density is offset by a higher price per pound (29¢/lb. for 12 x 40 vs. 26¢ /lb. for 8 x 30). These compensating factors result in a nearly equal price per cubic foot. The void fraction for each size carbon is nearly equal (see above) so that the "real" water contact time will be about the same even though the 50-minute contact time used in design is based on the empty vessel volume. Future changes relative costs and densities of carbon grades might change this considerably.

The comparison for the effect of particle size on economics is shown in Table 11, which shows that there is little difference between the operating costs and no significant difference in investment in the two cases. The carbon make-up cost and power requirement for the 12 x 40 mesh particle size are higher as expected. The assumption has been made in this comparison that particle size affects neither the required contact time nor regeneration loss. The effects of these variables will be examined in later sections.

Effect of Regeneration Loss on Economics
Cases 1A, 12 & 12A

This comparison shows the cost if the regeneration loss is 2, 5 or 10 percent. A loss of 2% represents the best performance which was obtained at the Wyandotte Chemical facility. A 5% loss represents the loss at Lake Tahoe while a 10% loss was the highest obtained at Pomona. Because the regeneration loss changes, the size of the fresh carbon storage bin changes, since this vessel must hold a six-month supply of make-up carbon. All other equipment remains the same in each case.

Losses can occur by attrition in the pipelines and screw conveyor, by attrition, decrepitation, gasification and burning in the regeneration furnace, or by thermal shock in the quench tank. Variables which may affect the regeneration loss are particle size and shape, type and degree of loading, amount of handling, regeneration severity and carbon base (such as bituminous coal, wood, coconut shell, etc.).

The comparison for the effect of regeneration loss on the costs is shown in Table 12, which shows that the regeneration loss has a substantial effect on the operating costs, while the investment is about the same for each case. The regeneration loss has an effect on the make-up carbon cost. The economics show that it is very important to minimize the regeneration loss and that there is incentive to make the regeneration process as efficient as possible - at least, try to reduce losses to 2 percent. The regeneration process should be studied carefully to pinpoint where the carbon loss actually occurs and to identify which variables affect the loss.

Effect of Particle Size and Regeneration Loss on Economics
Cases 1A, 8, 9 & 12

This comparison identifies the economic impact of the assumption that the observed difference in regeneration loss at Pomona (10% loss

TABLE 11

EFFECT OF PARTICLE SIZE ON ECONOMICS

Case No.	1A	8
Variable Tested	8 x 30 mesh Particle Size	12 x 40 mesh Particle Size
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	50
Particle Size, Mesh	8 x 30	12 x 40
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	6
Number of Trains	2	2
Investment, $M		
Concrete	121	121
Adsorbers	253	253
Tanks	5	5
Pumps	98	107
Special Equipment	36	36
Piping	158	158
Conveyors	4	4
Total Major Material	675	684
Plant Investment	1210	1219
Carbon	390 (1)	399 (2)
Fixed-Capital Investment	1600	1618
Operating Costs, ¢/M Gal.		
Make-up Carbon	1.09 (1)	1.22 (2)
Power @ 1¢/kwh	1.00	1.11
Backwash @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @ 50% Labor +1.485% PI/Yr.	1.02	1.02
Amortization @ 7.4% FCI/Yr.	3.24	3.28
Maintenance @ 5.75% PI/Yr.	1.91	1.92
Insurance @ 1% FCI/Yr.	0.44	0.44
Total Operating Cost	10.21	10.50

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(1) 8 x 30 mesh carbon @ 26¢/lb.
(2) 12 x 40 mesh carbon @ 29¢/lb.

TABLE 12

EFFECT OF REGENERATION LOSS ON ECONOMICS

Case No.	12	1A	12A
Variable Tested	2% Regen. Loss	5% Regen. Loss	10% Regen. Loss
Contactor Type	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7
Contact Time, Min.	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	2	5	10
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522
Vessel Size	25'6" I.D. x 30'	25'6" I.D. x 30'	25'6" I.D. 30'
Number of Vessels	6	6	6
Number of Trains	2	2	2
Investment, $M			
Concrete	121	121	121
Adsorbers	253	253	253
Tanks	4	5	11
Pumps	98	98	98
Special Equipment	36	36	36
Piping	158	158	158
Conveyors	4	4	4
Total Major Material	674	675	681
Plant Investment	1208	1210	1217
Carbon @ 26¢/lb.	390	390	390
Fixed-Capital Investment	1598	1600	1607
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	0.44	1.09	2.18
Power @ 1¢/kwh	1.00	1.00	1.00
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.02	1.02
Amortization @ 7.4% FCI/Yr.	3.24	3.24	3.26
Maintenance @ 5.75% PI/Yr.	1.90	1.91	1.92
Insurance @ 1% FCI/Yr.	0.44	0.44	0.44
Total Operating Cost	9.55	10.21	11.33

with 16 x 40 mesh carbon) and Lake Tahoe (5% loss with 8 x 30 mesh) are attributable to difference in particle size. As noted in the literature report[11], regeneration loss might be expected to be higher for the smaller size carbon since the smaller carbon is more active for adsorption (higher adsorption rate) and, therefore, should be more active for burning and gasification during regeneration.

Regeneration losses of 2 and 5 percent for 8 x 30 mesh carbon are compared with losses of 5 and 10 percent for 12 x 40 mesh. If regeneration loss is a function of particle size, then the larger particle should yield a lower loss. On this basis, the loss with 8 x 30 mesh carbon should be compared with a higher loss with the 12 x 40 mesh carbon. A loss of 2% represents the lowest reported regeneration loss obtained at the Wyandotte Chemical facility. A loss of 5% corresponds to the loss at Tahoe with 8 x 30 mesh carbon, while a loss of 10% was the highest obtained at Pomona with 16 x 40 mesh carbon. Table 13 shows the comparison for the effect of particle size and regeneration loss. From Table 13, if the 5% loss with 8 x 30 mesh carbon (Case 1A) is compared with the 10% loss and 12 x 40 mesh (Case 9), it is quite obvious that the 8 x 30 mesh carbon results in much lower costs. Even if the regeneration process could be improved so that the 8 x 30 mesh had a 2% loss (Case 12) and the 12 x 40 mesh had 5% loss (Case 8), the larger size carbon still is significantly less expensive. If particle size does affect regeneration loss in this way, then a larger size carbon should be used.

Effects of Particle Size and Contact Time on Economics
Cases 1A & 10A

This comparison identifies the economic result of the assumption that the contact time for a given removal is a function of particle size. The functionality is suggested in the literature report[12]. An extrapolation of Figure 16 in that report[13] was made to predict a 20% reduction in contact time at an 87% COD removal, when the particle size is changed from 8 x 30 to 12 x 40 mesh.

Case 1A (8 x 30 mesh carbon with 50-minute contact time) requires six 25'6" I.D. x 30' vessels while Case 10A (12 x 40 mesh with 40-minute contact time)requires six 25'6" I.D. x 25' vessels. As discussed in the section, "Effect of Particle Size," the carbon bed pressure drop is higher for the smaller size carbon (4 inches H_2O/ft. carbon for 8 x 30 mesh carbon vs. 8.5 in. H_2O/ft. for 12 x 40 mesh) and cost per pound of carbon is higher (26¢ for 8 x 30 vs. 29¢ for 12 x 40). However, the backwash rate of 12 x 40 mesh carbon is lower (10 GPM/ft.^{2} for 12 x 40 vs. 17 GPM/ft.^{2} for 8 x 30).

The economic comparison for these cases is shown in Table 14, which shows that the investment and operating costs are nearly the same for the case of 8 x 30 mesh carbon with 50-minute contact time and the case of 12 x 40 mesh with 40-minute contact time. If particle size affects only the required contact time, there is no difference in cost for the two particle sizes.

TABLE 13

EFFECT OF PARTICLE SIZE AND REGENERATION LOSS ON ECONOMICS

Case No.	12	1A	8	9
Variable Tested	8 x 30 Mesh 2% Regen.Loss	8 x 30 Mesh 5% Regen. Loss	12 x 40 Mesh 5% Regen. Loss	12 x 40 Mesh 10% Regen. Loss
Contactor Type	Downflow	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7	7
Contact Time, Min.	50	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	12 x 40	12 x 40
Regeneration Loss, %	2	5	5	10
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 30'	25'6"I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	6	6	6
Number of Trains	2	2	2	2
Investment, $M				
Concrete	121	121	121	121
Adsorbers	253	253	253	253
Tanks	4	5	5	11
Pumps	98	98	107	107
Special Equipment	36	36	36	36
Piping	158	158	158	158
Conveyors	4	4	4	4
Total Major Material	674	675	684	690
Plant Investment	1208	1210	1219	1225
Carbon	390 (1)	390 (1)	399 (2)	399 (2)
Fixed-Capital Investment	1598	1600	1618	1624
Operating Costs, ¢/M Gal.				
Make-up Carbon	0.44 (1)	1.09 (1)	1.22 (2)	2.44 (2)
Power @ 1¢/kwh	1.00	1.00	1.11	1.11
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.02	1.02	1.02
Amortization @ 7.4% FCI/Yr.	3.24	3.24	3.28	3.29
Maintenance @ 5.75% PI/Yr.	1.90	1.91	1.92	1.93
Insurance @ 1% FCI/Yr.	0.44	0.44	0.44	0.44
Total Operating Cost	9.55	10.21	10.50	11.74

(1) 8 x 30 mesh carbon @ 26¢/lb.
(2) 12 x 40 mesh carbon @ 29¢/lb.

TABLE 14

EFFECT OF PARTICLE SIZE AND CONTACT TIME ON ECONOMICS

Case No.	1A	10A
Variable Tested	8 x 30 Mesh Carbon 50 Minute Contact Time	12 x 40 Mesh Carbon 40 Minute Contact Time
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	40
Particle Size, Mesh	8 x 30	12 x 40
Regeneration Loss, %	5	5
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6" x I.D. x 25'
Number of Vessels	6	6
Number of Trains	2	2
Investment, $M		
Concrete	121	121
Adsorbers	253	246
Tanks	5	5
Pumps	98	103
Special Equipment	36	36
Piping	158	158
Conveyors	4	4
Total Major Material	675	673
Plant Investment	1210	1202
Carbon	390 (1)	322 (2)
Fixed-Capital Investment	1600	1524
Operating Costs, ¢/M Gal.		
Make-up Carbon	1.09 (1)	1.22 (2)
Power @ 1¢/kwh	1.00	1.06
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.03
Amortization @ 7.4% FCI/Yr.	3.24	3.09
Maintenance @ 5.75% PI/Yr.	1.91	1.94
Insurance @ 1% FCI/Yr.	0.44	0.42
Total Operating Cost	10.21	10.29

(1) 8 x 30 mesh carbon @ 26¢/lb.
(2) 12 x 40 mesh carbon @ 29¢/lb.

Effect of Particle Size, Contact Time and Regeneration Loss On Economics
Cases 1A, 10A, 11 & 12

This comparison identifies the economic consequence of the assumption that particle size affects both the contact time and regeneration loss. In this comparison, 8 x 30 mesh carbon with 50-minute contact time and 2% (Case 12) and 5% (Case 1A) regeneration loss is compared with 12 x 40 mesh with 40-minute contact time and 5% (Case 10A) and 10% (Case 11) regeneration loss.

The vessel requirements are six 25'6" I.D. x 30' vessels for Cases 1A and 12 and six 25'6" I.D. x 25' vessels for Cases 10 and 11. Other factors which change in these cases and which have been discussed in previous sections are pressure drop, fresh carbon storage and carbon cost.

The comparison showing the economic effect of particle size, contact time and regeneration loss is given in Table 15. By comparing Cases 1A and 11, if the use of 12 x 40 mesh carbon results in a 40-minute contact time and 10% regeneration loss, it is seen that the operating costs are substantially higher than for the larger size carbon. Similarly, if the regeneration losses are reduced to 2% and 5% for 8 x 30 (Case 12) and 12 x 40 (Case 10A), respectively, there is still a significant, though smaller, difference.

From the economics, it can be concluded that if particle size affects both required contact time and regeneration loss as suspected, then the larger size carbon is definitely preferred.

Effect of Carbon Capacity on Economics
Cases 1A, 13 & 14

This comparison identifies the economic effect if the carbon capacity is different from that assumed in the base case. Three capacities are examined:

1. 0.87 lb. COD/lb. carbon, the first regeneration cycle capacity at Pomona (Case 13).

2. 0.522 lb. COD/lb. carbon, the capacity predicted from Pomona data after the system reaches steady state with respect to number of regeneration cycles (Case 1A).

3. 0.25 lb. COD/lb. carbon, the capacity at Lake Tahoe (Case 14).

The effect of the number of regeneration cycles on capacity was taken from the literature report.[14] The steady state mentioned above was an extrapolated value read as 60% of the original capacity where the capacity appears to be leveling off. Since the capacity determines the regeneration rate, as the capacity decreases, the regeneration furnace size will increase. Also, since the regeneration loss increases with regeneration rate, the fresh carbon storage tank increases in size. The furnace size varies as follows for a 10-MGD plant:

TABLE 15

EFFECT OF PARTICLE SIZE, CONTACT TIME AND REGENERATION LOSS ON ECONOMICS

Case No.	1A	11	12	10A
Variable Tested	8 x 30 Mesh	12 x 40 Mesh	8 x 30 Mesh	12 x 40 Mesh
Contactor Type	Downflow	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7	7
Contact Time, Min.	50	40	50	40
Particle Size, Mesh	8 x 30	12 x 40	8 x 30	12 x 40
Regeneration Loss, %	5	10	2	5
Carbon Capacity, lb COD/lb C	0.522	0.522	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 25'	25'6"I.D. x 30'	25'6"I.D. x 25'
Number of Vessels	6	6	6	6
Number of Trains	2	2	2	2
Investment, $M				
Concrete	121	121	121	121
Adsorbers	253	246	253	246
Tanks	5	11	4	5
Pumps	98	103	98	103
Special Equipment	36	36	36	36
Piping	158	158	158	158
Conveyors	4	4	4	4
Total Major Material	675	679	674	673
Plant Investment	1210	1208	1208	1202
Carbon	390(1)	322(2)	390(1)	322(2)
Fixed-Capital Investment	1600	1530	1598	1524
Operating Costs, ¢/M Gal.				
Make-up Carbon	1.09(1)	2.44(2)	0.44(1)	1.22(2)
Power @ 1¢/kwh	1.00	1.06	1.00	1.06
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.02	1.02	1.03
Amortization @ 7.4% FCI/Yr.	3.24	3.10	3.24	3.09
Maintenance @ 5.75% PI/Yr.	1.91	1.90	1.90	1.94
Insurance @ 1% FCI/Yr.	0.44	0.42	0.44	0.42
Total Operating Cost	10.21	11.45	9.55	10.29

(1) 8 x 30 mesh carbon @ 26¢/lb.
(2) 12 x 40 mesh carbon @ 29¢/lb.

Capacity, lb. COD/lb. C	Regeneration Rate, lb./hr.	Furnace Size
0.25	735	8'6" O.D. - 6 Hearth
0.522	350	6' O.D. - 6 Hearth
0.87	212	39" I.D. - 10 Hearth

The comparison for the effect of capacity is shown in Table 16.

As can be seen in Table 16, the carbon capacity has a strong effect on the operating cost but very little effect on the investment. A spread in operating cost of 2¢/M gal. develops as the capacity changes from 0.25 to 0.87 lb. COD/lb. carbon. This provides sufficient economic incentive to try to maximize the capacity. This might be accomplished by changing the contactor type or operating conditions if it would increase the loading. Operating conditions might be found which would maximize the contribution of the biological action to the carbon's capacity. This might be accomplished by raising the water temperature and injecting air into the water.

Table 16 also shows how the operating costs will increase as the carbon is regenerated several times. Based on Pomona capacities, the operating cost when the carbon is new will be about 0.5¢/M gal. less than when it has been regenerated 10 times (Case 13 vs. Case 1A). However, the carbon probably would not degrade appreciably after 10 regeneration cycles, and costs would not continue to rise indefinitely. With the capacity at 0.522 lb. COD/lb. carbon, a vessel is regenerated about once every 6 months so that 10 regeneration cycles would take about 5 years. It is concluded that it is economically important to try to preserve the original capacity of the carbon during regeneration.

Effect of Contactor System Type on Economics
Cases 1A, 15 & 16

This comparison takes into account the differences in material, operating costs, and plant life for the downflow, upflow and gravity-flow contactors, making no allowance for differences in inherent efficiencies of the contactor. The gravity-flow contactor was compared with the downflow contactor at two plant sizes in a previous section of this report, but is also included in this comparison for completeness.

The downflow system is represented in this comparison by the base case (Case 1A). The upflow system concept is about the same as the downflow except for flow direction and the fact the suspended solids must be removed before the water enters the carbon bed. Since suspended solids cannot be backwashed off the bottom of a carbon bed, a sand and gravel filter is inserted before the carbon contactors to remove the suspended solids. Based on Pomona data[15], approximately 70%

TABLE 16

EFFECT OF CARBON CAPACITY ON ECONOMICS

Case No.	14	1A	13
Variable Tested	0.25 lb. COD/lb. C	0.522 lb. COD/lb. C	0.87 lb. COD/lb. C
Contactor Type	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7
Contact Time, Min.	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5
Carbon Capacity, lb COD/lb C	0.25	0.522	0.87
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	6	6
Number of Trains	2	2	2
Investment, $M			
Concrete	121	121	121
Adsorbers	253	253	253
Tanks	11	5	4
Pumps	98	98	98
Special Equipment	38	36	28
Piping	158	158	158
Conveyors	4	4	4
Total Major Material	683	675	666
Plant Investment	1219	1210	1199
Carbon @ 26¢/lb.	390	390	390
Fixed-Capital Investment	1609	1600	1589
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	2.29	1.09	0.66
Power @ 1¢/kwh	1.00	1.00	1.00
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.31	0.15	0.09
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.02	1.01
Amortization @ 7.4% FCI/Yr.	3.26	3.24	3.22
Maintenance @ 5.75% PI/Yr.	1.92	1.91	1.89
Insurance @ 1% FCI/Yr.	0.44	0.44	0.44
Total Operating Cost	11.60	10.21	9.67

of the total carbon capacity is for dissolved COD. On this basis, the capacity for the upflow contactor was chosen to be 0.348 lb. dissolved COD/lb. carbon. The COD concentration entering the plant is 60 mg/L and is reduced to 42 mg/L by the sand and gravel filter. The adsorbers do not have to have extra freeboard above the carbon bed to allow for expansion during backwash, since any backwashing would be downflow. This results in 25'6" I.D. x 25' vessel size. Otherwise, the flow sheet for the upflow system is the same as for the downflow system.

The gravity-flow system (Case 16) is the one described previously in the section "Effect of Gravity Flow Contactor at Two Plant Sizes on Economics." This system has 10 rectangular cross-section,single-stage vessels, each 20 ft. wide x 22 ft. long x 25 ft.-6 in. high. Each vessel has a 13 ft.-6 in. deep carbon bed, 2 ft. of sand and gravel under the carbon, with porous filter bottom under the gravel. A 10 ft. freeboard space above the carbon is provided for water rise as the bed pressure drop increases from suspended solids. Two separate troughs are provided for removal of backwash water and carbon for regeneration. The water velocity through the carbon bed is 2 GPM/ft.^{2} compared to 7 GPM/ft.^{2} for the downflow and upflow systems. Designing a single-stage, gravity-flow contactor for a higher velocity would result in deeper carbon beds and concrete walls which would be prohibitively thick. The contactor walls must be able to withstand the full water pressure since the adjacent contactor will be emptied for regeneration at times.

The cost comparison for the downflow, upflow and gravity-flow systems, is shown in Table 17. By comparing Cases 1A and 15, the upflow system is seen to be significantly more expensive than the downflow system. The additional cost is due to the sand and gravel filter needed for suspended solids removal. The higher operating cost is a reflection of the higher investment.

As observed in the section "Effect of Gravity Flow Contactor at Two Plant Sizes on Economics," the gravity-flow contactor is significantly less expensive than the downflow system (Case 16 vs. 1A). Investment savings are realized because concrete contactors are less expensive than steel vessels. In Table 17, the concrete cost includes field labor cost-the cost shown is for the concrete in place. Adsorber cost in Cases 1A and 15 is for material only. If the feed water and product pumps could be eliminated for a complete gravity-flow system, then the fixed-capital investment would be reduced by $71,000. The power and investment charges (at 15.6%/year) would be reduced by 0.55 and 0.30¢/M gal., respectively, for Case 16.

Once again, the major assumption made in the design of the gravity-flow system is that the required removal of COD is accomplished at 2 GPM/ft.^{2} with a 50-minute contact time. A carbon capacity of 0.522 lb. COD/lb. carbon in a single-stage contactor was also assumed. This latter assumption may not be such a large burden if the velocity is doubled and a two-stage contactor considered. This would give the same number of contactors as before but would require some additional piping and another pump; however, it would make the higher capacity possible.

TABLE 17

EFFECT OF CONTACTOR SYSTEM TYPE ON ECONOMICS

Case No.	1A	15	16
Variable Tested	Downflow	Upflow	Gravity
Contactor Type	Downflow	Upflow	Gravity
Velocity, GPM/ft^2	7	7	2
Contact Time, Min.	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5
Carbon Capacity, lb COD/lb C	0.522(1)	0.348(1)	0.522(1)
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 25'	20'w x 22'1 x 25'6"h
Number of Vessels	6	6	10
Number of Trains	2	2	8
Investment, $M			
Concrete	121	139	426
Adsorbers & Filters	253	363	-
Tanks	5	5	5
Pumps	98	107	116
Special Equipment	36	36	37
Piping	158	272	95
Conveyors	4	4	4
Total Major Material	675	926	683
Plant Investment	1210	1552	1064
Carbon @ 26¢/lb.	390	390	347
Fixed-Capital Investment	1600	1942	1411
Operating Costs, ¢/M Gal.			
Make-up Carbon @ 26¢/lb.	1.09	1.09	1.09
Power @ 1¢/kwh	1.00	1.06	0.64
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	1.16	0.96
Amortization @ 7.4% FCI/Yr.	3.24	3.94	2.24(2)
Maintenance @ 5.75% PI/Yr.	1.91	2.45	1.68
Insurance @ 1% FCI/Yr.	0.44	0.53	0.39
Total Operating Cost	10.21	11.74	8.51

(1) 70% of total carbon capacity is dissolved - 0.522 lb. total COD/lb. C - 0.348 lb. dissolved COD/lb. C

(2) 30-year plant life assumed - Amortization Rate = 5.8%/Yr.

Effect of Adsorbent Cost on Economics
Cases 1A & 17

This comparison shows the economic effect of changing the cost of the adsorbent, assuming that a lower cost adsorbent would give comparable process performance. This change affects the carbon inventory investment and the make-up carbon cost. An adsorbent cost range of 0 to 30¢/lb. is considered. When the adsorbent cost is zero, there is no point in regenerating the adsorbent, since it would simply be burned (for heating elsewhere) or discarded. For this situation, the regeneration system is not needed and the cost for this system is shown in Table 18 as Case 17. Since there is no regeneration furnace, the labor requirements is reduced from 1-1/4 to 1/4 men/shift. Also, there is no fuel required since all the fuel in Case 1A is used in the regeneration system.

A plot of the total operating cost against adsorbent cost is shown in Figure 3. As seen by the solid line on Figure 3, as the adsorbent cost drops from 30 to 0¢/lb., the operating cost decreases from 10.51 to 8.21¢/M gal. When the regeneration system is removed at 0¢/lb., the operating cost drops from 8.21 to 6.62¢/M gal. Thus, it is seen that adsorbent cost has a strong effect on the economics. It should be noted that there are other ways to draw the line in Figure 3. For example, the broken line on Figure 3 shows the total operating cost if the used adsorbent is discarded. It is seen at about 2.0¢/lb. adsorbent (net cost including possible fuel value credit) that the operating cost exceeds the cost when the adsorbent is regenerated. Therefore, at above 2.0¢/lb., it pays to regenerate the adsorbent.

Effect of Number of Contacting Stages on Economics
Cases 1A, 20, 20A & 20B

The purpose of this comparison is to discover if there is an optimum number of contacting stages. As the number of stages increases, the number of vessels and amount of piping will increase. But as the number of stages increases, so should the carbon capacity. That is, as the total bed depth (47 feet for 50-minute contact time at 7 GPM/$ft.^2$) is split into smaller segments, the average loading (capacity) of the carbon in the first contactor will increase. For example, consider the 2-stage and 4-stage system shown in Figure 4. It is assumed that breakthrough will occur in the effluent from the last vessel for both systems after the same length of time onstream. When breakthrough does occur, the first vessel in each system will be taken offstream for regeneration. In the 4-stage system, only 12 feet of carbon will be regenerated; in the 2-stage system, 24 feet will be regenerated. This 24 feet will have the same loading (capacity) as the first two vessels in the 4-stage case. But for the regeneration operation, an entire vessel must be taken offstream and regenerated at one time. Since the loading in the first 12 feet is higher than in the next 12 feet, the average loading for the entire 24 feet is lower than for the first 12 feet.

TABLE 18

EFFECT OF ADSORBENT COST ON ECONOMICS

Case No.	1A	17
Variable Tested	Regen. System	No Regen. System
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	50
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	-
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	6
Number of Trains	2	2
Investment, $M		
Concrete	121	121
Adsorbers	253	253
Tanks	5	5
Pumps	98	98
Special Equipment	36	4
Piping	158	158
Conveyors	4	4
Total Major Material	675	643
Plant Investment	1210	1171 (2)
Carbon	390 (1)	-
Fixed-Capital Investment	1600	1171
Operating Costs, ¢/M Gal.		
Make-up Carbon	1.09 (1)	- (2)
Power @ 1¢/kwh	1.00	0.99
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	-
Labor @ $3.50/Man-Hr.	1.05 (3)	0.21 (4)
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	0.58
Amortization @ 7.4% FCI/Yr.	3.24	2.37 (2)
Maintenance @ 5.75% PI/Yr.	1.91	1.84
Insurance @ 1% FCI/Yr.	0.44	0.32
Total Operating Cost	10.21	6.62

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(1) 8 x 30 mesh carbon @ 26¢/lb.
(2) 8 x 30 adsorbent at no cost
(3) 1-1/4 man/shift operating labor requirement
(4) 1/4 man/shift operating labor requirement

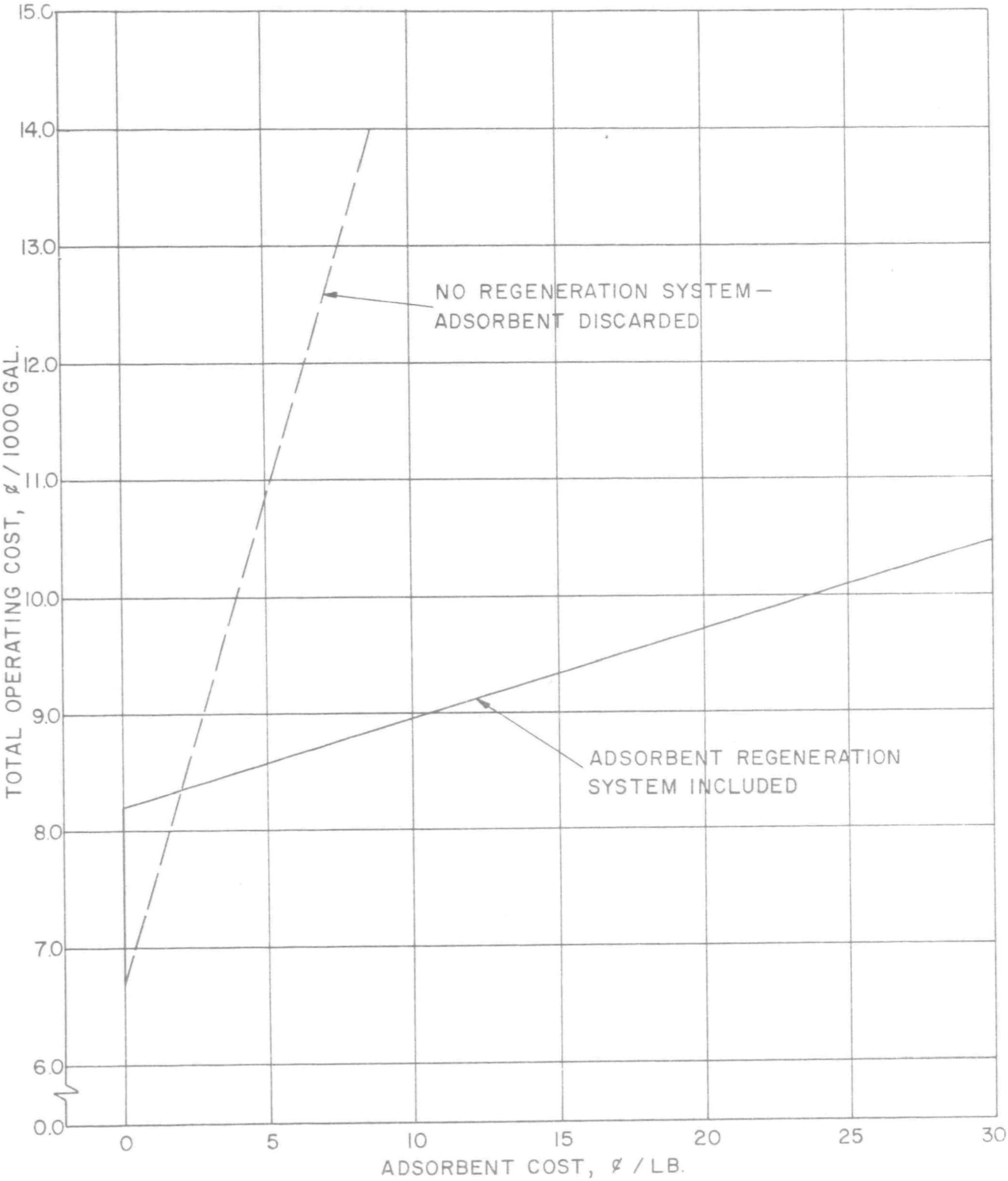
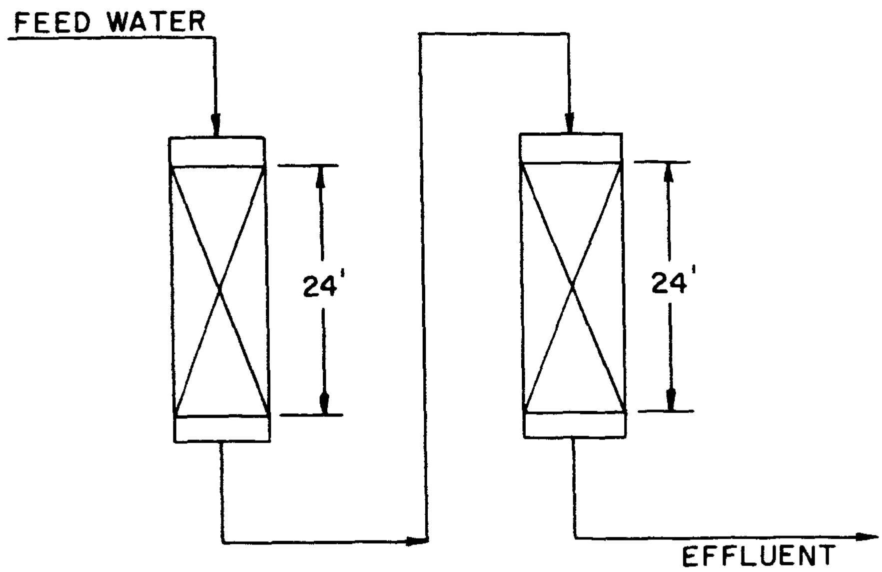
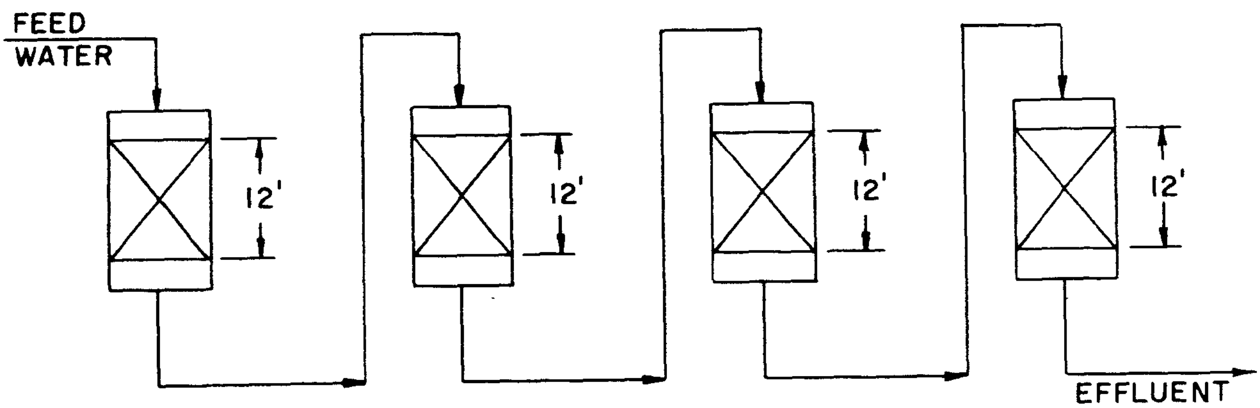


FIGURE 3

TOTAL OPERATING COST AS A FUNCTION OF ADSORBENT COST
TERTIARY WASTE WATER TREATMENT–GRANULAR ACTIVATED CARBON



TWO-STAGE SYSTEM



FOUR-STAGE SYSTEM

FIGURE 4

TWO-STAGE AND FOUR-STAGE SYSTEM DIAGRAM

The higher capacity for the 4-stage case results in a lower regeneration rate, smaller regeneration system and lower carbon loss. A carbon regeneration rate of 350 lb./hr. (same as base case) has been assumed for all cases for the purpose of regeneration system cost estimation. In the absence of a correlation of carbon capacity (loading) with number of contacting stages, a calculation in reverse order was attempted; i.e., an estimate of the approximate investment was made and then, with the difference in operating costs from the base case (2-stage) in hand, a back-calculation was made of the loading which must be achieved in order to pay for the extra equipment.

Considered in this comparison are one-, two-, three- and four-stage systems (Cases 20B, 1A, 20A and 20, respectively). Vessel requirements for the four cases are as follows:

Case	Stages	Vessel Size
20	4	34-13' I.D. x 16'
20A	3	26-13' I.D. x 21'
1A	2	6-25'6" I.D. x 30'
20B	1	6-18' I.D. x 61'

There are four trains of equipment in the single-stage system (even though the bed height-to-diameter ratio is much greater than one) in order to keep the amount of "idle" carbon at 1/5 or less of the total carbon inventory. (See section "Effect of Idle Carbon Inventory on Economics").

Shown in Table 19 are the costs for the 4-stage, 3-stage and 1-stage systems exclusive of the costs for make-up carbon or fuel for regeneration. The table shows that in the 4- and 3-stage cases (20 and 20A), the operating costs are already higher without make-up carbon and fuel than for the 2-stage system. This means that no matter what loading is attained in the 4- and 3-stage systems (as long as the loading is finite), costs will be higher for those systems than for a 2-stage system. When the single-stage system is examined (Case 20B), it is seen that its cost is lower (without make-up carbon and fuel included) than the 2-stage system. If it is then assumed that the single-stage system could achieve the same loading as the 2-stage system, the make-up carbon and fuel can be added in at 1.09 and 0.15¢/M gal., respectively, giving a total of 12.55¢ for 2-stage vs. 12.43¢ for single-stage, a slight difference. Of course, since a single-stage system could not possibly obtain as high a loading as a 2-stage system, the operating costs for the single-stage system exceed the costs for a two-stage system when loading is considered.

The conclusion is, therefore, that the two-stage contacting system represents a true optimum number of contacting stages. No further experimental work is needed to justify this conclusion. Loading data are required, however, in order to design the regeneration system for the two-stage system.

TABLE 19

EFFECT OF NUMBER OF CONTACTING STAGES ON ECONOMICS

Case No.	20	20A	1A	20B
Variable Tested	4-Stage Contactor	3-Stage Contactor	2-Stage Contactor	1-Stage Contactor
Contactor Type	Downflow	Downflow	Downflow	Downflow
Velocity, GPM/ft^2	7	7	7	7
Contact Time, Min.	50	50	50	50
Particle Size, Mesh	8 x 30	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	5	5	5	5
Carbon Capacity, lb COD/lb C	Not fixed	Not fixed	0.522	Not fixed
Vessel Size	13'I.D. x 16'	13'I.D. x 21'	25'6"I.D. x 30'	18'I.D. x 61'
Number of Vessels	34	26	6	6
Number of Trains	8	8	2	4
Investment, $M				
Concrete	156	141	121	116
Adsorbers	503	440	253	253
Tanks	5	5	5	5
Pumps	68	68	98	107
Special Equipment	44(1)	42(1)	36(1)	36(1)
Piping	602	367	158	142
Conveyors	4	4	4	4
Total Major Material	1382	1067	675	663
Plant Investment	2187	1682	1210	1177
Carbon @ 26¢/lb.	342	345	390	389
Fixed-Capital Investment	2529	2027	1600	1566
Operating Costs, ¢/M Gal.				
Make-up Carbon @ 26¢/lb.	-(2)	-(2)	1.09	-(2)
Power @ 1¢/kwh	1.07	1.03	1.00	1.03
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	-(2)	-(2)	0.15	-(2)
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.42	1.21	1.02	1.00
Amortization @ 7.4% FCI/Yr.	5.13	4.11	3.24	3.18
Maintenance @ 5.75% PI/Yr.	3.45	2.65	1.91	1.85
Insurance @ 1% FCI/Yr.	0.69	0.56	0.44	0.43
Total Operating Cost	13.12(2)	10.92(2)	10.21	8.85(2)

(1) Includes 6'O.D. x 6 Hearth Regeneration Furnace
(2) Make-up carbon and fuel gas costs not included - see text

Comparison of Pomona, Tahoe, Improved Downflow and Moving-Bed Systems Cases 5A, 18, 19 & 20

The purpose of these cases is to compare the Pomona and Lake Tahoe systems on as consistent bases as possible. Improved versions of the downflow and upflow systems with reasonable modifications for reducing the costs of treatment are included.

The system representing a scale-up of the Pomona pilot plant[18] from 0.3- to 10-MGD is presented as Case 20. This is the four-stage contacting system described in the preceding section, with the make-up carbon and fuel costs included and with 12 x 40 mesh carbon. The carbon used at Pomona is 16 x 40 mesh which is not commercially available. The capacity (0.522 lb. COD/lb.) is the same as that at Pomona after several regeneration cycles[5]. The contact time used in this case (50 minutes) is higher than that actually available at Pomona (41 minutes for 38 feet of carbon at 7 GPM/ft.^{2}). However, in order to keep the bed height-to-diameter ratio at about one (see discussion in "Design Bases"), with the total contact time of 41 minutes, the vessel diameter would have to be 10 feet. This diameter would require 13 trains of vessels for a total of 54 vessels (13 trains of 4 vessels each plus 2 spare vessels). In an effort to minimize the investment, the bed depth in each vessel was increased, so that 13 feet diameter vessels could be used, thus reducing the number of trains to 8 for a total of 34 vessels.

The Tahoe system[19] is represented by Case 18. An assumption made in this case is that 15-minute contact time with a moving carbon bed can contain the adsorption wave so that breakthrough does not occur. The actual Tahoe system operates on concentrations lower than the 42 mg/L of dissolved COD influent used in this study. The carbon capacity of 0.25 lb. dissolved COD/lb. carbon is the one calculated for Tahoe[16]. The liquid velocity at Tahoe is 7 GPM/ft.^{2}.

The improved downflow system (Case 5A) is the base case design (2-stage system) with the velocity increased to 10 GPM/ft.^{2} and contact time reduced to 35 minutes. This is the velocity-contact time functionality found in the literature report[9]. In addition, the regeneration loss is reduced to 2% which represents the performance at the Wyandotte Chemical facility.

The improved moving-bed system (Case 19) is a system with a 15-minute contact time as at Tahoe. The velocity has been increased to 10 GPM/ft.^{2} in order to reduce the number of trains of vessels. The carbon capacity has been increased from 0.25 to 0.348 lb. dissolved COD/lb. C which should be attainable since the carbon at Tahoe often had to be removed from the vessel before breakthrough due to plugging problems. The regeneration loss has been reduced to 2% based on Wyandotte experience.

No spare vessels have been included in Cases 18 and 19 since these are moving-bed systems where only part of the carbon bed is removed at a time for regeneration; at Tahoe, this has been done without taking the vessel offstream. The vessel sizes for these four cases is shown below:

Case	Vessel Size
20 - Pomona System	34-13' I.D. x 16'
18 - Tahoe System	8 - 13' I.D. x 15'
5A - Improved downflow	6 - 21' I.D. x 30'
19 - Improved upflow	6 - 13' I.D. x 22'

The economics for these four cases are shown in Table 20.

By comparing Cases 20 and 18, the Tahoe system is seen to be considerably less expensive than the Pomona system. The high costs in the Pomona system are due to the four-stage contactor system. On the other hand, when Cases 5A and 19 are compared, the two-stage downflow system compares favorably with the moving-bed system. As far as technical feasibility is concerned, the downflow system is probably on a much firmer experimental basis as are the assumptions which went into its calculation. The assumption in the moving-bed case - that the adsorption wave will be contained in the 15-minute contact time moving bed - seems rather optimistic. Operating data from Tahoe show that the COD concentration of 7 mg/L would not be achieved on the average[17]. For the downflow case, it remains to be proven that the higher velocity will reduce the required contact time.

Effect of In-Place Regeneration on Economics Cases 1A & 22

The purpose of this comparison is to discover if there is any economic incentive for developing a method for in-place regeneration. In this comparison, the base case (Case 1A) with the conventional multiple-hearth regeneration furnace is contrasted against a system where regeneration of the carbon occurs in the vessel (Case 22).

For the in-place regeneration system, it was assumed that the organic matter on the carbon would be cracked to lighter organic matter and/or volatilized by the time the carbon bed is heated to 1000°F. with 1200°F. superheated steam. For a vessel containing 300,000 lb. of carbon and with the gas velocity set at 0.25 ft./sec., a 30-hour period is required to heat the carbon to 1000°F. This calculation was made using principles of unsteady - state heat transfer. The calculation was repeated for a gas velocity of 1 fps with the result that the regeneration period was reduced to 7.5 hours; but this would increase

TABLE 20

COMPARISON OF POMONA, TAHOE, IMPROVED DOWNFLOW AND MOVING-BED SYSTEMS

Case No.	20	18	5A	19
Variable Tested	Pomona System	Tahoe System	Improved Downflow	Improved Moving-Bed
Contactor Type	Downflow	Moving-Bed	Downflow	Moving-Bed
Velocity, GPM/ft^2	7	7	10	10
Contact Time, Min.	50	15	35	15
Particle Size, Mesh	12 x 40	8 x 30	8 x 30	8 x 30
Regeneration Loss, %	10	5	2	2
Carbon Capacity, lb COD/lb C	0.522	0.25	0.522	0.348
Vessel Size	13'I.D. x 16'	13'I.D. x 15'	21'I.D. x 30'	13'I.D. x 22'
Number of Vessels	34	8	6	6
Number of Trains	8	8	2	6
Investment, $M				
Concrete	156	139	121	139
Adsorbers & Filters	503	211	253	194
Tanks	5	8	4	4
Pumps	68	107	92	107
Special Equipment	44	44	36	36
Piping	602	313	158	262
Conveyors	4	4	4	4
Total Major Material	1382	826	668	746
Plant Investment	2187	1291	1101	1173
Carbon	349 (1)	97 (2)	265 (2)	104 (2)
Fixed-Capital Investment	2536	1388	1366	1277
Operating Costs, ¢/M Gal.				
Make-up Carbon	2.44 (1)	1.56 (2)	0.44 (2)	0.44 (2)
Power @ 1¢/kwh	1.16	1.06	0.97	1.06
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.21	0.15	0.15
Labor @ $3.50/Man-Hr.	1.05	1.05	1.05	1.05
Overhead @ 50% Labor + 1.485% PI/Yr.	1.42	1.05	0.97	1.00
Amortization @ 7.4% FCI/Yr.	5.14	2.81	2.77	2.59
Maintenance @ 5.75% PI/Yr.	3.45	2.03	1.73	1.85
Insurance @ 1% FCI/Yr.	0.69	0.38	0.37	0.35
Total Operating Cost	15.81	10.46	8.76	8.80

(1) 12 x 40 Mesh carbon at 29¢/lb.
(2) 8 x 30 Mesh carbon at 26¢/lb.

the size of the steam generator by a factor of four. Pittsburgh Activated Carbon Company recommends that gas velocities be kept between 0.25 and 1 fps so the 0.25 fps case was selected for the estimate. The total amount of steam used for in-place regeneration is 2/3 lb. steam/lb. carbon - actually less than the amount of steam required in the furnace operation. A nine-inch refractory brick lining is provided to protect the vessel shell from this temperature. Also, since there is no regeneration furnace where there must be a carefully controlled combustion, the operating labor can be reduced from 1.25 to 0.25 men/shift. An afterburner is included on the effluent gas stream for pollution control.

The cost comparison for the effect of in-place regeneration is shown in Table 21. As seen in Table 21, the cost for in-place regeneration is about 1¢/M gal. less than the furnace system while the investment is almost $300,000 higher. Reductions in the in-place regeneration operating costs came about from savings in items related to labor (1.22¢/M gal.) and make-up carbon (1.09¢/M gal.). Part of these savings are offset by higher investment charges.

In summary, there is incentive to develop a regeneration process where the carbon loss by attrition and regeneration loss is reduced and where there is no need for a closely supervised, controlled-combustion furnace. A completely automated regeneration system would cost more, but about $300,000 could be spent for controls before the investment charges would offset the savings in labor.

TABLE 21

EFFECT OF IN-PLACE REGENERATION ON ECONOMICS

Case No.	1A	22
Variable Tested	Furnace Regeneration	In-Place Regeneration
Contactor Type	Downflow	Downflow
Velocity, GPM/ft^2	7	7
Contact Time, Min.	50	50
Particle Size, Mesh	8 x 30	8 x 30
Regeneration Loss, %	5	0
Carbon Capacity, lb COD/lb C	0.522	0.522
Vessel Size	25'6"I.D. x 30'	25'6"I.D. x 30'
Number of Vessels	6	5
Number of Trains	2	2

Investment, $M

Concrete	121	123
Adsorbers	253	366
Tanks	5	-
Pumps	98	100
Special Equipment	36	40
Piping	158	162
Conveyors	4	-
Total Major Material	675	791
Plant Investment	1210	1488
Carbon @ 26¢/lb.	390	390
Fixed-Capital Investment	1600	1878

Operating Costs, ¢/M Gal.

Make-up Carbon @ 26¢/lb.	1.09	-
Power @ 1¢/kwh	1.00	0.98
Backwash Water @ 6¢/M gal.	0.31	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.08
Labor @ $3.50/Man-Hr.	1.05 (1)	0.21 (2)
Overhead @ 50% Labor + 1.485% PI/Yr.	1.02	0.71
Amortization @ 7.4% FCI/Yr.	3.24	3.81
Maintenance @ 5.75% PI/Yr.	1.91	2.34
Insurance @ 1% FCI/Yr.	0.44	0.51
Total Operating Cost	10.21	8.95

(1) 1-1/4 Man/shift operating labor
(2) 1/4 Man/shift operating labor

CONCLUSIONS AND RECOMMENDATIONS

Based on economics and arguments given in the "Economics" Section, the following conclusions and recommendations can be drawn. The conclusions and recommendations are listed in the order of their importance. Some of the recommendations are the same as presented in the literature report[1] and are now being carried out in FWPCA programs.

1. The gravity-flow system offers the savings of 2¢/M gal. over the downflow system and a savings of 3¢/M gal. over the upflow system. Gravity-flow is less expensive because concrete vessels are less expensive than steel vessels. An assumption made for the gravity-flow case and which would have to be proven, is that a 50-minute contact time will give the same COD removal at 2 GPM/ft^2 as at 7 GPM/ft^2 and that the carbon loading (capacity) will be the same for single- and two-stage contactors. Recommendation: The performance of a single-stage, gravity-flow contactor should be determined.

2. Under the same design conditions, the downflow system is less expensive than the upflow system by over 1¢/M gal. Upflow is more expensive because sand filters are required to remove suspended solids. Recommendation: No new data are needed to confirm this conclusion.

3. A two-stage contactor is an optimum pressurized vessel system as compared to single-, three- and four-stage contactors. The costs are seen to be higher for 3- and 4-stages due to increased vessel and piping costs than for 2-stages, no matter what carbon capacity is assumed. When capacity is considered in the single-stage system, this system is also more expensive than the two-stage system. Recommendation: A two-stage, downflow contacting system containing 8 x 30 mesh carbon should be operated at a large scale for extended periods in order to determine degree of removal and carbon capacity.

4. The comparison between the Pomona and Tahoe systems as reported in the literature[18,19] finds the Tahoe system to be substantially less costly (about 5c/M gal.). Pomona is more expensive due to the four-stage system used. An assumption made for the Tahoe moving-bed system is that the 15-minute contact time will contain the adsorption wave which is doubtful based on Tahoe operating data. Recommendation: No new data are needed to confirm this conclusion.

5. When reasonable improvements are made to the downflow (Pomona-type) and moving-bed (Tahoe-type) systems, the costs are about equal. However, the downflow system is felt to be on a firmer experimental basis. Recommendation: No additional work is needed on upflow systems.

6. Substantial savings can be realized if a less expensive adsorbent can be found. A reduction in adsorbent cost of 15¢/lb would reduce carbon treating costs by about 1.15¢/M gal. Also, up to a new price of about 2.0¢/lb ($40 per ton) for the adsorbent, it would be cheaper to discard the adsorbent rather than regenerate it. Recommendation: A search for lower cost adsorbents should be made.

7. Significant savings can be realized if the regeneration loss can be reduced to 2%. Operating costs can be reduced by alomst 2¢/M gal. when the regeneration loss is cut from 10% (Pomona) to 2% (Wyandotte Chemical). Recommendation: An effort should be made to pinpoint where the carbon loss occurs and to understand the mechanism of regeneration so that regeneration losses can be minimized.

8. A high carbon capacity for COD is to be preferred in order to reduce costs associated with the regeneration system. Operating costs can be cut by 2¢/M gal. when the capacity is increased from 0.25 lb COD/lb C (Tahoe) to 0.87 lb COD/lb (Pomona first regeneration cycle). Recommendation: Schemes should be examined which would increase the allowable loading (capacity) of the carbon. This might include enhancement of the biological action by injecting air (oxygen) into the wastewater.

9. Costs would be significantly lower if the closely controlled combustion in the regeneration furnace could be eliminated. If a highly automated furnace operation could be accomplished, a savings in labor cost of 1.2¢/M gal. is possible. In-place regeneration offers the same advantages plus the possibility of reducing the attrition and regeneration losses. Recommendation: Complete automatic control of the regeneration furnace or elimination of the furnace regeneration system should be investigated.

10. Handling of surges in feed flow rate with a surge basin rather than with extra vessels is less expensive by nearly 1.3¢/M gal. with a considerable reduction in investment. Recommendation: No new data are required to confirm this conclusion.

11. Even though costs are not significantly different between shop fabrication and field erection of vessels, the field erection case offers a simpler flow sheet. Recommendation: No new data are required to confirm this conclusion.

12. Costs are not affected substantially if the "idle" carbon inventory is between 10% and 20% of total carbon inventory. Recommendation: No new data are needed to confirm this conclusion.

13. Substantial economics of scale are seen when the plant size is increased from 1- to 10-MGD. This reduction is on the order of 20¢/M gal. due mainly to reduced labor cost. When the plant size is increased from 10- to 100-MGD, the operating cost is reduced by only 3.5¢/M gal. The reduction is smaller at the large plant size because there are several trains of vessels required. Recommendation: No new data are required.

14. If it is assumed that particle size affects the regeneration loss and required contact time, then the case with a larger particle size, longer contact time and lower regeneration loss is less costly. Reductions of up to 2¢/M gal. are possible when 8 x 30 mesh carbon is used rather than 12 x 40 mesh. If particle size affects only the required contact time, then there is no difference in costs between 8 x 30 and 12 x 40 mesh carbon. Recommendation: The functionality between carbon particle size and regeneration loss should be determined.

15. When the required contact time is assumed to decrease with increasing velocity, then the high velocity-low contact time case is less costly by almost 3¢/M gal. Recommendation: The relation between velocity and required contact time should be determined.

16. When it is assumed that velocity does not affect the required contact time, then the 7 and 10 GPM/ft^2 cases are not significantly different in cost. However, the 4 GPM/ft^2 case is more expensive by 1.7¢/M gal. Recommendation: No new data are needed to confirm this conclusion.

17. There is little difference in operating costs and no significant difference in investment when just the particle size is changed from 8 x 30 to 12 x 40 mesh carbon. If changing the particle size over this range affects only the pressure drop and carbon cost, then either size carbon could be used without any penalty in costs. Recommendation: No new data are required to confirm this conclusion.

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APPENDIX A

PROCEDURE FOR CALCULATION OF OPERATING COSTS

The following is the basis and rationale used in our recommendation for calculation of the operating costs for tertiary waste water treatment using granular activated carbon. Economic comparisons, so long as they are on consistent bases, should be valid no matter what particular percentages for maintenance and overhead items are chosen for comparison purposes. No other comparisons should be made unless care is taken to make sure that the cases in question are on consistent bases.

It is felt that the activated carbon system is more like a typical chemical plant than a water or sewage treatment plant as the activated carbon system has some corrosion problems and has a furnace where the combustion must be carefully controlled. Operating labor, for example, is based on experience in a clay treating plant for taste, odor and color removal from hydrocarbons (oils and waxes) which Kellogg has designed and built. The activated carbon system is quite similar to this process.

The recommended procedure is based on several sources among which are the Office of Saline Water, the Office of Coal Research as well as well-known texts on engineering economics. A more precise estimation of the maintenance and overhead charges can be determined only from actual costs obtained from large-scale plants. Maintenance charges for individual pieces of equipment can be more accurately estimated from an extensive examination of literature and vendor recommendations and, especially, plant operating records. However, maintenance costs are quite variable depending on operating personnel quality and amount of inspection and preventive maintenance. In the activated carbon process, maintenance costs could be quite high if a more corrosion-resistant lining for the vessels is not found. At the current state of knowledge of this process, we feel that these recommendations represent the most realistic estimate of the operating costs and are consistent with chemical industry and utility practice.

A review has been made of several different general procedures for calculation of maintenance and overhead costs of operating a chemical plant. Among the sources examined are the Office of Saline Water[1], Office of Coal Research[2], Kellogg internal practice[3], Aries and Newton[4], and Peters[5]. The following are our recommendations for calculation of the operating costs for our economic evaluation.

Maintenance Labor	-	3% Plant Investment/year
Maintenance Materials	-	2% Plant Investment/year
Supplies	-	15% of Maintenance Labor & Materials
Payroll Overhead	-	15% of Operating plus Maintenance Labor
General & Administrative Overhead	-	30% of Operating plus Maintenance Labor plus Payroll Overhead
Amortization	-	7.4% Fixed-Capital Investment/year (20-year life with 4% interest/year)
Insurance	-	1% Fixed-Capital Investment/year
Taxes	-	None (normally 2% Fixed-Capital Investment/year)

The first three items equal 5.75% of Plant Investment/year while the next two items total to 50% of Operating Labor plus 1.485% of Plant Investment/year. The total of the items which are dependent upon the Plant Investment is 15.6%/year while that dependent on Carbon Investment only is an additional 8.4%/year.

Table A shows the comparison of the various calculation procedures and, for illustration, gives the operating cost for Case 1A. The operating cost can vary from 8.22¢/M gal. to 12.64 ¢/M gal., depending on the method used. The maintenance, supplies and labor costs are somewhat higher than in conventional waste water treating (primary and secondary) but are nearer to costs experienced in typical chemical plants which carbon treating more closely resembles. On the other hand, the overhead items are typical for the type of governmental agency which would be operating this plant. A more accurate estimation of the maintenance and overhead costs will be available only after a sustained operation of some of the large-scale units such as at Lake Tahoe or Piscataway, Maryland.

The items labeled supplies, payroll overhead and general and administrative overhead include the following:

- Supplies include instrument charts, lubricants, janitor supplies, test chemicals, and similar supplies which cannot be considered as raw or maintenance materials.

- Payroll overhead includes pensions, paid vacations, group insurance, disability pay, social security, and unemployment taxes.

- General and administrative overhead includes costs for supervision, medical services, general plant maintenance and overhead, safety services, packaging, restaurant and recreation facilities, salvage services, control laboratories, property protection, plant superintendence, warehouse and storage facilities, special employee benefits, purchasing, engineering, executive and clerical wages, office supplies, communications, advertising and consultant fees.

Other operating costs such as power, fuel gas, and backwash water are charged at unit costs which are typical for these items. The backwash water is taken from the secondary treatment plant and returned to the intake for primary treatment. The unit cost for producing secondary effluent varies from 3¢ to 9¢/1000 gallons depending on plant size.

A sufficient initial charge of activated carbon must be bought at the time that the plant is built to fill the vessels. This carbon wears out, not unlike the vessels which contain it, and therefore, must be depreciated. For cost accounting purposes, the initial charge of carbon is amortized over the same period as the plant. The initial charge of carbon eventually gets entirely replaced by the make-up carbon which is charged as an operating cost. The inventory of make-up carbon is not included in the Fixed-Capital Investment since it is bought a little at a time and can be sold back to the supplier if not used. No salvage value has been assumed for the used carbon which remains at the end of

TABLE A

COMPARISON OF PROCEDURES FOR OPERATING COST CALCULATION

METHOD ITEM RECOMMENDATION	OSW[1]	OCR[2]	MWK TO AEC[3]	ARIES & NEWTON[4]	PETERS[5]	POMONA[6]	MWK RECOMMENDATION
a. Operating Labor	--	--	--	--	--	1 man-day/day	1-1/4 men/shift
b. Maintenance Labor	0.5% PI/yr	3% PI/yr	2.8% PI/yr	3% PI/yr	3% PI/yr	2/3 man-day/day	3% PI/yr
c. Maintenance Materials	} 0.5% PI/yr	} 15% (b)	1.2% PI/yr	3% PI/yr	4% PI/yr	2% PI/yr	2% PI/yr
d. Supplies			15% (b + c)	15% (b + c)	10% (b + c)	25% (c)	15% (b + c)
e. Payroll Overhead	15% (a + b)	10% (a + f)	20% (a + b + f)	15-20% (a)	w/(h)	25% (a + b + f + g)	15% (a + b)
f. Supervision		10% (a)	15% (a)	10-25% (a)	10-20% (a)	5% (a + b + g)	
g. Laboratory Labor	} 30% (a + b + e)	} 50% (a + b + c + d + f)	} 50% (a + b + c + d + f)	} 50-100% (a)	} 65-85% (a + b + c + f)	1/3 man-day/day	} 30% (a + b + e)
h. General Overhead						} 15% (a + b + f +g)	
i. Plant Insurance	1% FCI/yr	} 3% FCI/yr	} 3% FCI/yr	1% FCI/yr	0.4-1% FCI/yr		1% FCI/yr
j. Taxes	1% FCI/yr			1-2% FCI/yr	1-4% FCI/yr	none	none

CASE 1A OPERATING COSTS*							
Make-up Carbon @ 26¢/lb	1.09	1.09	1.09	1.09	1.09	1.09	1.09
Power @ 1¢/kwh	1.00	1.00	1.00	1.00	1.00	1.00**	1.00
Backwash Water @ 6¢/M gal.	0.31	0.31	0.31	0.31	0.31	0.05	0.31
Fuel Gas @ 25¢/MM Btu	0.15	0.15	0.15	0.15	0.15	0.15	0.15
Amortization @ 7.4% FCI/yr	3.24	3.24	3.24	3.24	3.24	3.24	3.24
Plant Insurance @ 1% FCI/yr	0.44	0.44	0.44	0.44	0.44	w/ovhd	0.44
Taxes - None	--	--	--	--	--	--	--
a. Operating Labor @ $3.50/man-hour	1.05	1.05	1.05	1.05	1.05	0.84	1.05
b. Maintenance Labor	0.17	0.99	0.93	0.99	0.99	0.56	0.99
c. Maintenance Materials	} 0.17	} 0.15	0.40	0.99	1.33	0.66	0.66
d. Supplies			0.20	0.30	0.23	0.17	0.25
e. Payroll Overhead	0.18	0.12	0.43	0.18	w/ovhd	0.42	0.31
f. Supervision		0.11	0.16	0.18	0.16	0.08	
g. Laboratory Labor	} 0.42	} 1.15	} 2.74	} 0.79	} 2.65	0.28	} 0.71
h. General Overhead						0.25	
TOTAL OPERATING COST, ¢/M gal.	8.22	9.80	12.14	10.67	12.64	8.79	10.20

METHOD
ITEM RECOMMENDATION
a. Operating Labor
b. Maintenance Labor
c. Maintenance Materials
d. Supplies
e. Payroll Overhead
f. Supervision
g. Laboratory Labor
h. General Overhead
i. Plant Insurance
j. Taxes

CASE 1A OPERATING COSTS*
Make-up Carbon @ 26¢/lb
Power @ 1¢/kwh
Backwash Water @ 6¢/M gal.
Fuel Gas @ 25¢/MM Btu
Amortization @ 7.4% FCI/yr
Plant Insurance @ 1% FCI/yr
Taxes - None
a. Operating Labor @ $3.50/man-hour
b. Maintenance Labor
c. Maintenance Materials
d. Supplies
e. Payroll Overhead
f. Supervision
g. Laboratory Labor
h. General Overhead

* Case 1A

Plant Investment (PI)	1,210,000
Carbon @ 26¢/lb	390,000
Fixed Capital Investment (FCI)	1,600,000

** Backwash Water @ 1¢/M gal.

the plant's life even though the carbon could be used in another waste treatment plant. It is felt that the used carbon could not be sold back to the supplier or to anyone else at the virgin carbon cost.

The cost estimates in this report have been made on the basis of a Gulf Coast, U.S.A. plant location. Included in the plant investment estimate are such items as foundations, concrete structures, vessels, tanks, drums, structural steel, pumps, compressors, piping, wiring, switchgear, lighting, instruments, paint, insulation, conveyors, cranes, special equipment (regeneration furnace system, eductors, water softener, air pollution control system, steam generator) and construction labor. Also included in the plant investment, are non-material items such as all-risk insurance (liability, accident and loss insurance during construction), field office administration, supervision and expenses, home office procurement, engineering and scheduling, and contractor's fee.

Not included in the plant investment are working capital, interest charges during construction, local taxes, freight and duties. Also, land has not been included in the investment, especially since even a 100-MGD gravity-flow plant would occupy only about two acres which at $30,000 an acre is still a very small percentage of the total cost (about 1%). Ordinarily, the cost of land is not included in the investment since it is the responsibility of the owner and not the contractor to buy the land for the plant site.

REFERENCES FOR APPENDIX A

1. Office of Saline Water, "A Standardized Procedure for Estimating Costs of Saline Water Conversion," March, 1956.

2. The M. W. Kellogg Company, Report RED-68-1173 to the Office of Coal Research, Contract No. 14-01-0001-380, September 1, 1968, p. 123.

3. The M. W. Kellogg Co., Report RD-62-952 to the U.S. Atomic Energy Commission, Contract No. AT (30-1)-3009 (NYO 10,301), November 30, 1962, p. 167.

4. Aries, R. S. and R. D. Newton, "Chemical Engineering Cost Estimation," McGraw-Hill Book Company, Inc., New York (1955) pp. 162-182.

5. Peters, M. S. "Plant Design and Economics for Chemical Engineers," McGraw-Hill Book Company, Inc., New York (1958) pp. 108-9, 113.

6. Letter from C. W. Carry, Los Angeles County Sanitation Districts to A. N. Masse, Federal Water Pollution Control Administration, October 31, 1968.