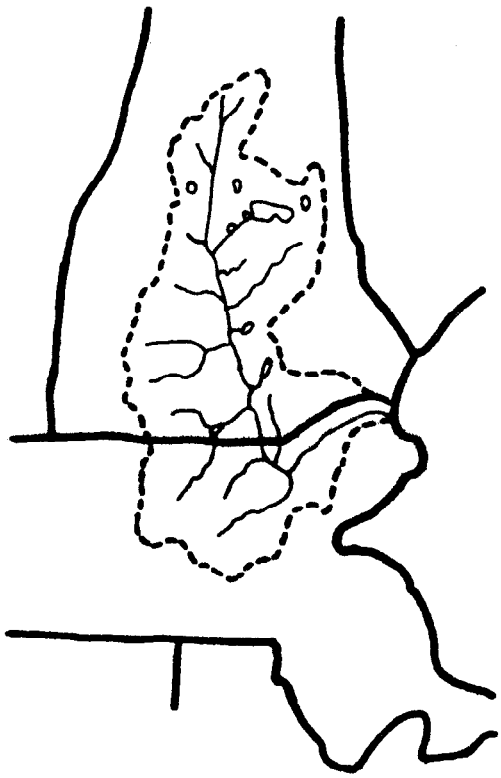
THE MERRIMACK:

DESIGNS FOR A CLEAN RIVER



CONSULTANT'S LAND AND WATER DISPOSAL ASSESSMENTS

ANNEX A

AUGUST 1971

ANNEX A
to
THE MERRIMACK WASTEWATER FEASIBILITY STUDY

Following is a list of data prepared by private consultants for use in this feasibility study:

Annex No. A - Land and Water Disposal Consultants

Title of Report	Author
Preliminary Report on Physical Characteristics Relative to Waste Management by Land Application in The Merrimack Basin and the Boston Metropolitan Area	Prof. James E. Hackett
Soils Systems for Renovation of Wastewater	T. D. Hinseley Ph.D.
Industrial Waste Effects of the Wastewater Management Program for the Merrimack River Basin in Massachusetts and New Hampshire.	Alfred M. Tenny, TENCO Hydro/Aeosyences, Inc.
Land Disposal Alternatives	Mr. Wayne Cowlishaw
Storage and Transportation Systems	Dr. William Bauer
Reverse Osmosis	Franklin J. Agardy, Ph.D.
Tertiary Treatment by Infiltration in the Merrimack River Basin	E. Robert Baumann, Ph.D.
Ultimate Disposal	Joseph F. Malina, Jr. Ph.D.
Report on Waste Treatment in the Merrimack River Basin by Ion Exchange	R. L. Sanks, Ph.D.
Report on Adsorption for the Merrimack Basin Pilot Wastewater Management Program	Prof. W. J. Weber, Jr.
Wastewater Treatment by Coagulation in the Merrimack River Basin	Prof. Mark Tenney
Preliminary Examination Instream Aeration Alternative of theMerrimack River	Water Resources Institute Rutgers University William Whipple Jr. and Shaw L. Yu

PRELIMINARY REPORT ON PHYSICAL CHARACTERISTICS RELATIVE TO WASTE MANAGEMENT BY LAND APPLICATION IN THE MERRIMAC BASIN AND THE BOSTON METROPOLITAN REGION

James E. Hackett

PROBLEM DEFINITION

The principle objective of this study is to evaluate the regional characteristics with regard to potentials for land application of municipal wastewater from the Boston Metropolitan area and from municipal systems within the Merrimac basin. Current wastewater volume anticipated is approximately 540 MGD with about 400 MGD from Metropolitan Boston and about 140 MGD from basin communities. For planning purposes a 1990 volume of 881 MGD is used. Land area requirements to manage wastewater will vary depending on rates of application, length of growing season, location of treatment cells and holding basins, and amount of peripheral off-site location control areas. The actual irrigation area would be approximately 126,000 acres (143 acres/MGD (Muskegon Experience) x 881 MGD (design flow) = 126,000 acres). Additional area would be required for the aerated treatment cells and the storage basins.

MERRIMAC BASIN

Area of the drainage basin is about 5,010 square miles (3,206,400 acres). Above Lowell, the basin contains 4,635 square miles (2,966,400 acres). Therefore if a minimum of 126,000 acres of irrigated land is required, slightly less than 4 per cent of the basin will be irrigated with treated wastewater.

Owing to a lack of detailed geologic and soils data over large portions of the basin, specific site identification cannot be accomplished at this stage of the investigation. However, regional characteristics inferred from the informa-

tion provided allows for the definition of several major site areas within which certain potentials for land application of wastewater exist. The general physical characteristics of the site areas also permit identification of factors that bear on management concerns within each site area.

Regional Characteristics

For purposes of identification of major site areas, the basin has been subdivided into four broadly defined terrains:

1. Mountainous (steep slope) terrain.
2. Intermediate slope terrain.
3. Ice-contact deposit terrain.
4. Outwash and alluvium terrain.

Owing to the lack of subsurface control, to a deficiency in precise surficial mapping over large parts of the basin and to a lack of adequate correlation between soil association units and geological units, boundaries between terrains cannot be considered precise but are broadly generalized and largely interpretive.

General description of the terrains and the area of their occurrences are summarized in the following table:

Terrain	General Characteristics	Occurrence
1. Mountainous	Steep slopes generally in excess of 15%; mountainous terrain with thin till cover (thickness generally less than 20 feet). Extensively forested. Little subsurface storage. Surface drainage fairly well developed. Application rates restricted to soil moisture deficiency in growing season. Surface storage capabilities locally available.	Northern and western parts of Merrimac Basin in New Hampshire.

	Intermediate slope	Area of intermediate slope between the mountainous terrain (terrain 1) and the more level slopes of terrains 3 and 4. Merges with terrain 1. Till and ice-contact cover generally less than 20 feet in thickness but locally thicker. Topography is irregular and drainage is poorly developed. Contains rapid variation of soil type and locally high saturation levels. Subsurface storage capacity is variable but generally low. Application restricted to soil moisture deficiency in growing season. Irregular application patterns would be required. Land use is a mixture of small farms and forest land. Fragipan occurs commonly at shallow depth. A subdivision of this terrain is those areas containing thicker drift mainly of till. This subregion is also characterized by restricted application rates and limited areal extent for application patterns. It differs in that there is some opportunity for land modification to facilitate application of wastewater.	Widely distributed throughout the basin, intermediate between terrain 1 and 3 in the seaboard lowlands and between 1 and 4 in the Northern New England Upland and Southern New England Upland.
3.	**Ice-contact deposit**	Ice-contact deposits at surface to depths in excess of 20 to 30 feet extending to thicknesses in excess of 100 feet. Topography is generally irregular with locally steep slopes. Extensively developed in the Lowell area south of the Merrimac River (USGS WSP 1669-Y Plate 2). Material is included in soil association 39 (Gloucester, etc.) and is generally permeable. Subsurface storage potential ranges from low to high depending on local saturation level. Opportunities for ground water management exist but are affected by variations in texture. Potential application rates range from medium to high. Areas underlain by this terrain are extensively farmed.	Present in extensive deposits in Seaboard Lowlands south of Lowell and possibly south of Manchester adjacent to and underlying the broad deposits of outwash.
4.	**Outwash and Alluvium**	Relatively level topography with good internal drainage except in alluvial areas where saturation levels are near surface. Subsurface storage potential generally high, especially in higher level terraces. Opportunity for water level management programs by well and tiling. These opportunities are more limited in alluvial lowlands adjacent to the streams. Local	Along the Merrimac River and its major tributaries--a) north of Franklin, b) south and west of Concord, c) south of Manchester to Nashua, d) vicinity of Lowell.

presence of Lake Clays, especially in the vicinity of Concord, may limit infiltration characteristics. Potential application rates range from medium to high. Outwash and alluvium is underlain by permeable ice-contact deposits in the vicinity of Lowell. Similar conditions may exist south of Manchester to near the state line. Land in this terrain is extensively farmed.

Site Areas for Land Application

Considering the regional characteristics of the Merrimac Basin in terms of terrainal units, several site areas with various potentials for land application of wastewater are identified. The locations of the site areas are shown on the accompanying map.

Site Area 1: Lowell Area

Extent and Characteristics. Extensive areas of ice-contact deposits are present in the Lowell Area adjacent to and south of the Merrimac River. These deposits arrain thicknesses in excess of 100 feet, are generally coarse in texture and permeable. They are closely associated with, and in places underlie, the outwash and alluvial deposits. Information on the distribution and subsurface extent of the ice-contact deposits in the Lowell Area are described and displayed in USGS WSP 1669-Y. Within the report area some 20 to 30 square miles are estimated to be underlain by these materials.

The presence of ice-contact deposits in association with outwash and alluvium indicate good potential for management of ground water levels by wells or tiling. However, the ice-contact deposits are marked by broad range in textural characteristics and hydrologic interconnection within the deposits is likely to be variable. The Lowell Area is situated within the Seaboard Lowland physiographic province. The extent of ice-contact deposits beyond the region mapped in

the USGS report is not known; however, it is possible that these conditions occur widely within the Seaboard Lowlands physiographic province. Land surface is generally irregular but good internal drainage reduces the incidence of ponding or marsh development.

Management Factors. The nearness of the Lowell Area to the Boston Metropolitan area suggests that large parts of this area either have been developed or are very high in land value. This would limit extensive use of this region for irrigation land application purposes. Use of the infiltration and a[illegible] (flooding basin) method could be considered in the Lowell Area provided techniques to operate the facilities satisfactorily under severe winter conditions can be devised. Assuming that with this method an application rate of 150 feet per year can be accomplished as compared to 8.5 feet per year by the irrigation method, the area requirements are reduced from 143 acres per MGD to 8.2 acres per MGD. Considering 90 wastewater plus storm water volumes of 131 MGD from the Lowell-Haverhill area, a land area of approximately 1075 acres would be required.

Site Area 2: Manchester Area

Extent and Characteristics. A broad and extensive area of outwash occurs along the Merrimac River Valley and its major tributaries from about Manchester to Nashua, a distance of approximately 20 miles. Within this area, the outwash deposits attain widths of between two to six miles. Outwash deposits averaging about a mile in width also extend up the major tributaries on the west of the Merrimac River, a distance of some 10 to 20 miles. Approximately 100 square miles of outwash and associated ice-contact deposits are estimated to be present in the area. The thicknesses and character of the outwash deposits of this region are not specifically known, but they are likely to be permeable and to have sufficient depth be suited for water-level management activities. The Manchester area occurs

within a unit of the Seaboard Lowland province as does the Lowell site and the potential in the Manchester to Nashua area for associated deposits of ice-contact materials similar in character and relationships to occurrences in the Lowell area should be carefully investigated.

Management Factors. The extent of development in the outwash deposits between Manchester and Nashua is not known but is believed to be not as intense as in the Lowell Area, nor are land values likely to be as high. Land surface is likely to be generally level on the outwash terraces, but the associated ice-contact deposits, if present, are likely to have more irregular surfaces. This region displays good potential for land application of wastewater at rates required for the irrigation or infiltration-seepage methods of land application. The underlying permeable material provides the opportunity for successful management of water levels by wells or tiling.

Site Area 3: Concord Area

Extent and Characteristics. There are extensive outwash and alluvial deposits present in the vicinity of Concord north of Manchester. These deposits with a width of two or four miles extend for a distance of approximately 15 miles along the Merrimac River and for a distance of approximately 10 miles up the major tributary of the Merrimac, the Contoocook River. An estimated total of between 60 and 70 square miles of outwash deposits occur here. The broad area of outwash in the Concord area is separated from that of the Manchester area by a narrowing of the Merrimac River Valley at the lower end of the Northern New England Upland as the river passes from that upland onto the Seaboard Lowland in which Manchester occurs. The outwash deposits in the vicinity of Concord are known to contain lake clay sediments in the upper part of the outwash materials.

here present, these lake clays would tend to impede the infiltration of water into the lower levels of the outwash materials. The extent of the lake clay within the outwash is not known but several exposures of the clay have been noted in this region.

The restriction in width of the Merrimac River Valley between Concord and Manchester, the presence of lake clays in the Concord region and the general directions of ice movement through this area suggest the possibility of a presence of a buried or partially buried valley, between these two major outwash areas. If present, the drift filled valley could have a bedrock floor at an elevation of 100 to 500 feet above sea level, and could therefore, contain a drift till in excess of 350 feet. If the fill of the projected valley consists largely of outwash and/or ice-contact deposits, additional opportunities for land application of wastewater might be present.

Management Factors. The outwash and alluvial deposits in the vicinity of Concord have relatively level land surfaces which are conducive to land application practices. The presence of lake clay at shallow depth, however, may seriously restrict application rates. The community of Concord occupies a large portion of the outwash area along the Merrimac River and availability of disposal areas is likely to be very restricted.

Site Area 4: Upland Mountain Area

Extent and Characteristics. A large part of the Northern and Western areas of the basin are occupied by steep-sloped and heavily forested mountains. The crystalline bedrock in the mountain areas is overlain by a relatively thin blanket of drift materials, generally less than 20 feet in thickness. This terrain is included largely in soil association areas 8, 9, 10 and 13 and, therefore, volves nearly 40% of the basin area or approximately 2,000 square miles. The

drift consists largely of till and associated ice-contact deposits and is, therefore, variable in texture. Because of the dense nature of the crystalline rocks and the thinness of drift cover, subsurface storage capacities are generally limited. However, because of the nature of this terrain, facilities for surface storage are widely available.

The thinness of the cover material and the steep slopes stimulate rapid runoff from this area and during the growing season the forest soils frequently are droughty. During long periods of drought the small quantities of subsurface water held in storage are soon exhausted. Therefore, although rates of precipitation and runoff are high for this region, soil moisture deficiencies may exist for significant portions of the growing season. Because of the extremely limited subsurface storage capacities and anticipated variation in texture of cover materials, acceptable application rates of wastewater are likely to be significantly less than two inches per week. On the other hand, the broad expanse of area and the undeveloped nature of this terrain indicates that there is some potential for handling large volumes of wastewater in this site area even at very low application rates.

Management Factors. Due to the steep slopes and rugged terrain, accessibility would be a major concern. Because of extended periods when application of wastewater would be restricted, surface storage would have to be provided, but opportunities for such storage are widely available. Application would have to be conducted under conditions of steep slopes and heavy forest cover. Application rates would have to be carefully adjusted to soil moisture conditions to guard against excessive runoff as a consequence of application operations. An additional concern is that with soil moisture held at capacity or near capacity, a larger portion of the natural precipitation would occur as direct runoff. More

frequent flooding in adjacent lower lying areas could thereby result. Soils on the steep slopes maintained at or near soil moisture capacity attain saturated conditions more rapidly during rain events. During periods of heavy rains or rains of long duration, conditions of instability due to excessively saturated soils on steep slopes could occur.

Site Area 5: Hillsboro Area

Extent and Characteristics. A narrow but relatively long area of outwash occurs in the upper part of the Contoocook basin in the vicinity of Hillsboro. The character and thickness of the outwash and alluvium in this area is not known; however, if permeable, areas may be available for wastewater application, particularly in the head waters portion of this drainage system. The potential of this area for land application should receive serious consideration in the event that wastewater is transported for disposal in the adjacent mountain areas. The length of outwash and alluvial deposits along this drainage is approximately 20 miles; however, the width of the deposit seldom exceeds two miles and the areal extent of the deposits probably does not exceed 20 square miles.

Management Factors. A principal concern is the distance of transport of wastewater to this relatively isolated area of small dimension. If, however, wastewater is transported for application in the broad mountain areas to the north, this site area is likely to have greater significance.

Site Area 6: Northern Tributaries Area

Extent and Characteristics. North of Franklin, along the Merrimac River and its tributaries, outwash and alluvium with an average width of approximately one mile extends some 35 miles along the drainage valleys. The character and thickness of the outwash in the northern tributaries valleys is not known. How-

ever, if permeable, and if adequate subsurface storage exists, this region of some 35 square miles could have a significant potential for wastewater application activities.

Management Factors. The relatively narrow width of the valley areas indicates that wastewater applied on these outwash deposits would move directly to the associated surface streams. Subsurface detention time would not be as long in this site area as in those areas of broader outwash deposits to the south. The limited areal extent in relation to the distance for transport of wastes from producing areas is also a constraining factor.

Other Site Areas

The remainder of the basin apparently is occupied by a relatively thin cover of glacial material generally less than 20 feet in thickness over a dense crystalline bedrock. The drift cover is extremely variable in texture ranging from silty clay till to coarse textured ice-contact deposits. The drift surface is highly irregular with poorly developed drainage. The thinness of the cover and the variable texture of the material severly limits the opportunity for land application over sizeable tracts. In these areas there is some possibility for treatment of wastewater by overland flow methods. Operations of this type may be suitable for handling the wastes generated by the Winnipesaukee River area. South and north of Laconia, tracts of land totalling approximately 80 square miles are present where land surfaces appear to be more regular than in other portions of the region. If materials underlying this area are too thin to provide for any sizeable subsurface storage for the treatment of wastes from the Lake Winnipesaukee area, the overland runoff method of treatment could be given consideration.

General Conclusions

The existence in the Merrimac Basin of environments suited to accomodate the large volume of wastewater generated both in the Boston Metropolitan area and in the basin is questionable. Sizeable areas (square miles in extent) relatively consistent in character, with infiltration capacity to accomodate application rates in excess of three inches per week, with sufficient subsurface storage potential and with transmissibilities adequate to permit management of ground-water levels by wells or tiles are severely restricted in occurrence and are generally limited to regions underlain by outwash and alluvium or by extensive ice-contact deposits generally in association with outwash and alluvium.

Following are those site areas in which such conditions are most likely to occur and the estimated extent of these requisite conditions within site areas;

Site Area	Estimated Maximum Extent of Suitable Deposits	*Probable Usable Area
Lowell Area	30-40 square miles	10-15% (3-6 sq. miles)
Manchester Area	100 square miles	25-50% (25-50 sq. miles)
Concord Area	60-70 square miles	10-25% (6-17 sq. miles)
Hillsboro Area	20 square miles	15-25% (3-5 sq. miles)
Northern Tributaries Area	35 square miles	10-20% (3-7 sq. miles)
TOTALS	245-265 square miles	40-85 sq. miles

Assuming a land area requirement of 197 square miles to handle a wastewater load of 881 MGD at an application rate of three inches per week of growing season, it appears that the area demands for wastewater application is very likely to exceed the estimated total extent of area probably suited for such application programs. Area requirements for the infiltration-seepage method would be about

*Estimates based on consideration of extent of development, relative land values, distribution pattern of deposits, and probably thickness restrictions.

11 square miles and could be accomodated within the basin.

There is some possibility that with intensive use of all available acreage best suited for wastewater application, sufficient acreage could be utilized to handle through irrigation methods the wastewater load generated within the Merrimac Basin. Land area requirements for this loading are estimated to be in the order of 69 square miles for the irrigation method or about 3 square miles for the infiltration-seepage method.

There is also some possibility for considering the application of wastewater to the steep-sloped, heavily forested mountain terrain that occupies about 2,000 square miles within the Merrimac Basin. Land area requirements, assuming an application rate of one inch per week during the growing season, would be in the order of 1057 square miles to accomodate the total wastewater load of 881 MGD. More than one-half of the land area of the mountain terrain would be required. This alternative introduces considerable question of the economic and physical feasibility of such an operation.

From the practical standpoint, it appears that disposal of wastewater from the Boston Metropolitan area by the irrigation method cannot be accomodated in the Merrimac Basin without extensive use of the mountain terrain and alternative areas within the region should be searched for if only this method is to be used. It can also be concluded that disposal by the irrigation method of wastewater generated in the Merrimac Basin would likely require intensive land use management in the areas of the more suitable terrains. These conclusions suggest the following alternatives with regard to the use of land application of wastewater in the Merrimac Basin:

1. Use of irrigation methods of land application for wastewater generated in the basin above Lowell. This should result in improved water quality in the Merrimac River at Lowell where water withdrawals to serve the Boston area are anticipated. The Lowell-Haverhill and Boston Metropolitan Area wastewaters could be accomodated by land disposal at suitable sites in the coastal lowlands of Southwestern Maine and Southeastern Massachusetts.

2. Use of irrigation and infiltration-seepage methods of land application for wastewater generated within the Merrimac Basin including the Lowell-Haverhill area. Application areas would be located at and upstream of Lowell. The most promising areas are that of the outwash and possible associated ice-contact deposits in the vicinity of Lowell between Manchester and Nashua. More detailed physical studies would need to be conducted to assess the feasibility of this alternative. The major benefits would be improvement in water quality of the Merrimac River throughout its extent and increased flows at the projected water supply withdrawal point at Lowell. Wastewater from the Boston Metropolitan area can be handled as in alternative one. In the event of the use of improved treatment at Boston for ocean disposal of effluent, the sludge produced might be applied as a soil builder on the more permeable outwash and ice-contact deposits.

3. Use of infiltration-seepage method of land application in the Merrimac Basin of wastewater generated in the Lowell-Haverhill area and

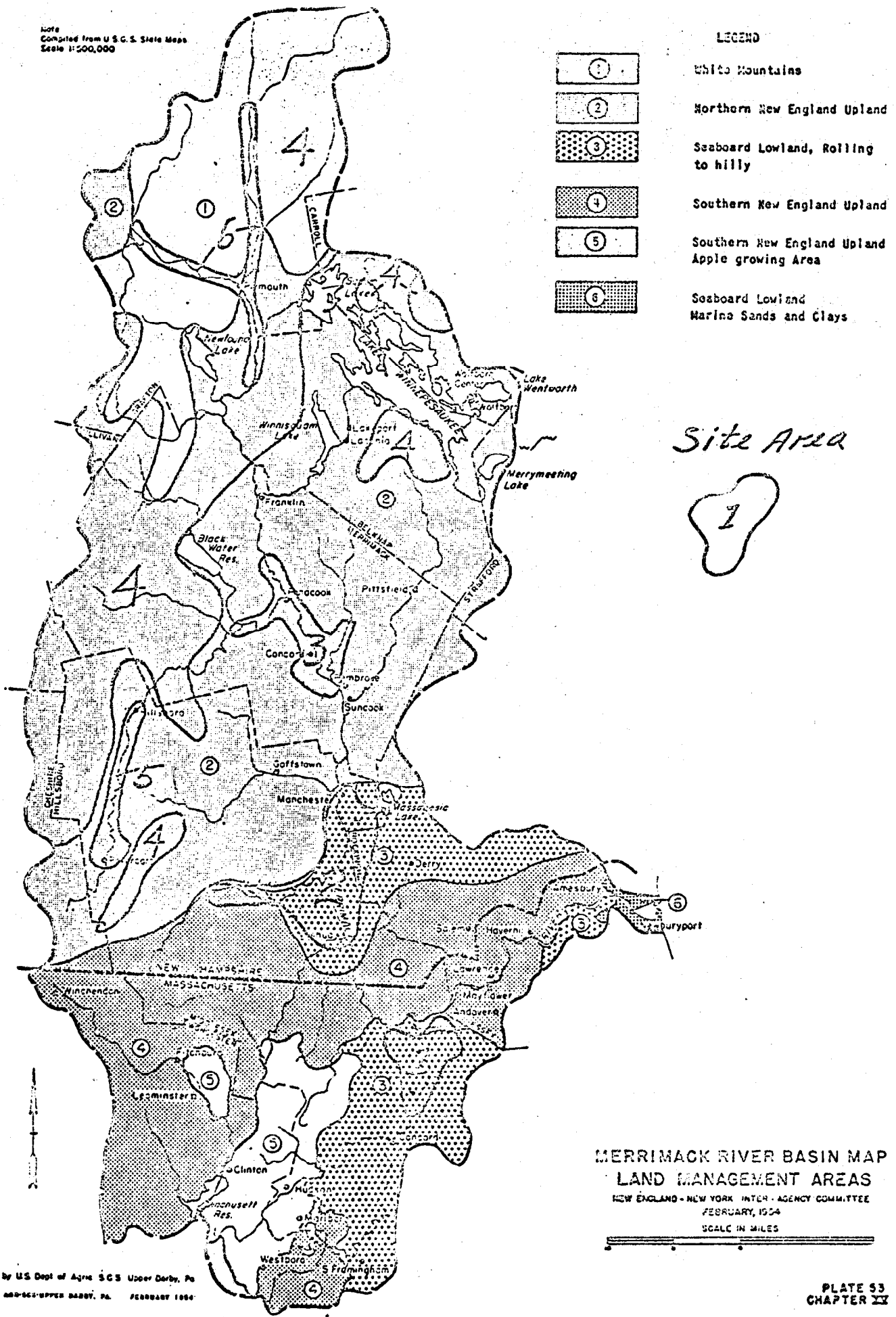
by the Boston Metropolitan area. Wastewater plus storm water flow for year 1990 would be 881 MGD (750 MGD + 131 MGD). Land area requirements would be 7,225 acres (8.2 x 881) or 11.3 square miles. Adequate acreage is believed to be present in the vicinity of Lowell and in the Manchester-Nashua area. The remainder of the wastewater plus storm water generated within the basin could be accomodated locally by whatever land application method is most appropriate to local site conditions.

The above conclusions regarding the potentials for land application within the Merrimac Basin indicates the desirability of locating suitable site areas outside of the basin to accomodate Boston or Boston plus Lowell-Haverhill waste water flows. Land area requirements by the irrigation method would be 107,250 acres (750 x 143) for Boston and 126,000 acres (881 x 143) for Boston plus Lowell-Haverhill. The land area requirements by the infiltration-seepage method would be 6,150 acres and 7,225 acres, respectively.

SOUTHWESTERN MAINE

The region of southwestern Maine from Portland to the Piscataqua River is an area of relatively thick deposits of glacial drift and glacial marine sediment. The glacial drift consists of glacial till, ice-contact deposits and outwash. The glacial marine sediments consist chiefly of fine-textured silts and clays. Topography ranges from relatively level to gently rolling on the glacial marine and outwash deposits to irregular with locally steep slopes on the ice-contact materials. The glacial till terrain is irregular and hilly.

Land application of wastewater is best adopted to those areas underlain



MERRIMACK RIVER BASIN MAP
LAND MANAGEMENT AREAS
NEW ENGLAND - NEW YORK INTER - AGENCY COMMITTEE
FEBRUARY, 1954
SCALE IN MILES

PLATE 53
CHAPTER XX

by ten feet or more of the more permeable outwash and ice-contact deposits. As shown on U. S. Geological Survey Hydrologic Atlas HA-76, broad tracts of land in this part of Maine are underlain by these materials. Generally the ice-contact deposits border the outwash deposits on the west. The outwash deposits range from about ten feet to more than forty feet in thickness generally increasing in thickness and becoming coarser textured in the direction of the ice-contact materials. The broad tracts of outwash and ice-contact materials are separated by till plains These materials are underlain by and interbedded with glacial marine sediments which are widely exposed in the eastern parts of the region. Drainage valleys commonly intersect the contact between the outwash and the marine sediments exposing the marine deposits in the valleys.

Site Areas

An extensive area of outwash and ice-contact deposit is present west of Kennebunk. West of the north-south toll road there are at least 25 square miles of permeable terrain suited for irrigation application. To the north of this tract, additional deposits of at least equal extent are present. South of the Kennebunk tract, several smaller tracts in the order of 2 to 3 square miles in extent are present. It is estimated that there are at least 50 square miles of land area suited for irrigation application of wastewater in the southwestern Maine region. Portions of these areas especially in the thicker ice-contact deposits would also be suited for the infiltration-seepage method of application.

SOUTHEASTERN MASSACHUSETTS

Regional Characteristics

South of Boston, the glacial drift is generally thicker than in the regions to the north and northwest. In this area the glacial drift is composed mainly

of glacial till and outwash. Ice-contact deposits are less prevalent and of smaller extent than in the Merrimac Basin or southwestern Maine regions. Land surface is less rugged with broad areas of relatively level or gently sloping land interspersed with more hilly tracts. Marsh lands are extensive and in central Plymouth County are locally used for cranberry production.

In the Cape Cod area thick deposits of glacial outwash occur as a frontal apron to a well-defined terminal moraine that forms the northern side of the cape. Permeable outwash exceeding a thickness of 100 feet have been reported in wells drilled in this area.

To the north, in central Plymouth County, outwash and minor ice-contact deposits produced by apparently stagnant ice conditions underlie a large part of the lower more level land areas including the marsh areas. These outwash areas are locally interrupted by uplands and hills of tighter glacial till. These conditions appear to be characteristic of a large part of the Sarragansett Basin, a physiographic low lowland extending northeastward of Providence, Rhode Island into central Plymouth County. Thicknesses of the outwash deposits are not known but are anticipated to extend to forty feet or more below land surface. Ground water saturation levels are expected to be relatively high in these outwash materials.

Site Areas

At the western end of Cape Cod, at and adjacent to Otis Air Base, there is an extensive area with a relatively level land surface underlain by permeable outwash attaining thicknesses of at least 100 feet. Approximately 40 square miles of contigious area well suited for irrigation applications are estimated to be present at this location. Additional acreages of suitable materials are present to the eastward in the central part of the Cape south of the terminal

moraine. An additional 25 square miles of area is estimated to be present in this region.

In central Plymouth County, north of the Cape Cod area, broad outwash areas interspersed with uplands and hills provide additional acreage suited to irrigation application. Approximately 100 square miles of area within which about 25 square miles has been estimated as having adequate thicknesses of outwash suited to land application methods is present in central Plymouth County. Additional areas of similar conditions are probably present westward within the Sarroganset Basin to Providence. Of significant concern in this region, however, is the heavy forest cover that is present throughout the area.

General Conclusions

There apparently is adequate land area suited for irrigation application of wastewater available in southwestern Maine and southeastern Massachusetts to handle the projected 1990 domestic sewage and storm water flows from the Boston area or from the Boston plus Lowell-Haverhill areas. Land area requirements are in the order of 165 square miles for the Boston flows and 240 square miles for the Boston plus Lowell-Haverhill flows.

Area and site requirements for five-month storage require further design consideration before the physical feasibility for storage facilities can be evaluated.

PRELIMINARY CONSULTANT REPORT:

SOIL SYSTEMS FOR RENOVATION OF WASTEWATER

BY

T. D. HINSLEY, PhD
Soil Specialist
University of Illinois

1971

SOIL SYSTEMS FOR RENOVATION OF WASTEWATER

In this report irrigation of crops at rates just sufficient to satisfy their need for additional water is not considered. In the humid eastern half of the United States, areas of droughty soils are exceedingly scarce in comparison with the volumes of wastewater to be utilized for crop irrigation. Even on the most droughty soils, it is unlikely that yields of general farm crops would be increased by applications of additional water in amounts greater than 12 to 14 inches per year during the growing season.

To maintain an application rate of four inches per week in a humid area where average precipitation is sufficient to meet 70 to 75 percent of the maximum evapotranspirational demand during the growing season will likely require removals of BOD from wastewater to levels of less than 40 PPM. Greater BOD contents in wastewater would likely result in a severe oxygen dificiency to growing plants, reducing their ability to utilize nutrients. Prior treatment of wastewater to decrease BOD levels will produce a concomitant decrease in the total nitrogen content to perhaps not more than 30 PPM. During the growing season about 80 inches of water would be applied at a rate of four inches per week. If two-thirds of the nitrogen is in an inorganic form, about 350 pound per acre of readily available nitrogen will be supplied to the crop. Considering yields in the neighborhood of 6 tons of dry weight per acre for a grass or silage crop containing 3 percent nitrogen the uptake and removal of nitrogen from the soil system

by harvesting the crop will be about equal to the soluble nitrogen supplied during the growing season. Further removals of nitrogen from additional application of wastewaters during dormant vegetative periods will depend on gaseous losses and rates of nitrogen immobilization as residues and organics contained in wastewater are assimilated by microorganisms. Immobilization of a part of the total nitrogen and the production of increasing amounts of stable organic matter would be expected, where amounts supplied exceed amounts removed by cropping. But at some point in time it would also be expected that equilibrium conditions between rates of immobilization and mineralization of nitrogen to reduce the soil system's effectiveness in removing nitrogen from percolating wastewater to that attainable by crop removal.

While the nitrogen contents of the wastewater may limit the amount that can be applied, phosphorus contents may eventually determine the usefulness of a particular site for wastewater renovation. Although, soils generally have a tremendous capacity for fixing phosphorus, it is not insatiable.

Cole and Olsen (1959) reported maximum surface adsorption values of 9.74, 16.9, and 24.6 miligrams of phosphorus per 100 grams of a loam, clay loam, and clay soil respectively. In light of these values it seems safe to assume that medium to coarse sandy soils may have the capacity to fix 20,000 pound of phosphorus per acre foot depth before they lose their ability to effect a

90 percent reduction in the phosphorus content of percolating wastewater, Assuming further that the phosphorus content of the wastewater is 20 PPM and that 4 inches can be applied for 34 weeks, the total application of phosphorus would amount to approximately 600 pounds per acre per year. Neglecting the 20 to 25 pounds per acre that might be remove by cropping it would require 30 to 35 years to saturate a soil one foot deep with phosphorus. Therefore, if the drainage tiles are four feet deep the useful life of the soil system should not be less than 125 years.

Because of their lower specific surfaces, the solubility of phosphorus increases more in sandy than clay textured soils with the same treatment rates. Furthermore, the solubility of phosphorus with application of additional amounts depend on the initial saturation of the soil system. The slow reactions of phosphorus with iron, aluminum, and calcium in soils to form minerals with very low solubilities may extend the life of the soil system. Like nitrogen some of the phosphorus will be incorporated into new microbial cells and later released by mineralization. Thus, to predict the useful life of a soil system to extract sufficient quantities of phosphorus from wastewater is a difficult task. When the cation exchange capacity of soils containing expanding type clay minerals is satisfied to the extent of 10 to 15 percent by sodium ions, permeability may be considerably decreased as soil aggregation is destroyed by

the dispersion of colloidal particles. Coarse textured sandy soils are less susceptible to the deterioration of the physical condition by the accumulation of adsorbed sodium as compared to medium and fine textured soils. However, some of the suspended solids in wastewater are clay minerals which after being filtered out of several feet of wastewater by the first few centimeters of the soil surface may exert a much larger influence on infiltration rates than would be expected on the basis of their total content. The tendency of sodium to replace divalent cations on exchange position during the application of wastewater can be estimated by determining the proportion of sodium to calcium plus magnesium present, together with the total salt content. The SAR (sodium adsorption ratio) is defined as $Na^{+}/\sqrt{Ca^{++}+Mg^{++}/2}$ expressed as miliequivalents per liter. Although some stone fruits may be injured, most general crops will not be affected by SAR values of less than 8. Wastewaters having SAR values of 18 may present few problems for use on crop plants if the total dissolved solids are low. With high rates of water applications and in the absence of the salt concentrating action of high evaporation rates at the surface of soils, it seems probable that the higher SAR value would be applicable in the humid eastern part of the country if total dissolved solids are not greater than 500 mg/l.

Since boron contents in sludges from waste treatment plants are exceedingly low practical all of it contained in the influent must remain in soluble forms in the effluent. Boron may need to

be given special consideration relative to other trace elements in wastewater because concentrations in soil solutions greater than 1 PPM are toxic to some crop plants. Boron is an essential plant nutrient and is often absorbed by plants in relatively large amounts (20 to 60 PPM) but will damage all plants at soil solution concentrations greater than 5 PPM. Boron may exist in soil as boric acid, calcium or magnesium borates, and as constituents of silicate minerals. Boron is readily taken out of solution by illite clay minerals and fixed in a nonexchangeable form. It is believed that boron replaces aluminum in the tetrahedral positions of illites (Harder 1963).

Like boron the content of molybdenum in sludges from waste treatment plants is always very low or not detectable, indicating that most of the element entering the plant is probably discharged in soluble forms with effluents. Molybdenum exist in soils in the solution phase, as an exchangeable trioxide anion, and as constitutent of minerals and organic matter. When the soil solution contains more than 0.01 PPM of molybdenum, some forage plants can accumulate the element to concentrations greater than the 5 PPM dry weight content generally considered to be a maximum safe limit for feeding cattle. Molybdenum is one of the few elements which can be accumulated in plants to detrimental levels for feeding without producing toxic symptoms in the growing plants.

Other trace elements in wastewater can be maintained in sparingly soluble form limiting their absorption by growing plants,

by management practices that insure an aerobic rooting zone and soil pH values of not less than 6.5. Nevertheless, it is prudent to demand the removal at the source of the wastewater those chemical element species that exceed the tolerance levels recommended for irrigation waters in the U.S. EPA publication entitled "Water Quality Criteria".

REFERENCES

1. Cole, C. V. and Olsen, S. R. 1959, Phosphorus solubility in calcareous soils II. Effects of exchangeable phosphorus and soil texture on phosphorus solubility. Soil Sci. Soc. Amer. Proc. 23, 119-121.

2. Harder, H. 1963, To what extent is boron a marine index element Fortschr. Geol.Rheinld Westf. 10, 239-253.

PRELIMINARY CONSULTANT REPORT:

INDUSTRIAL WASTE EFFECTS

ON THE

WASTEWATER MANAGEMENT PROGRAM

FOR THE

MERRIMACK RIVER BASIN

IN

MASSACHUSETTS & NEW HAMPSHIRE

BY:

ALFRED M. TENNY

TENCO HYDRO/AEROSCIENCES, INCORPORATED

INDUSTRIAL WASTE EFFECTS
ON THE
WASTEWATER MANAGEMENT PROGRAM
FOR THE
MERRIMACK RIVER BASIN
IN
MASSACHUSETTS & NEW HAMPSHIRE

SCOPE: The supplementary report will be limited to a review of possible sources and types of industrial wastewaters. The effect of these wastes on the effluent quality from any proposed land disposal system [LDS] will also be considered, together with means of controlling adverse effects due to these pollutants. The reduction of nutrient levels and soluble B.O.D. will be considered as part of the industrial waste problem.

BKGD: Assuming adequate land is available, a fact which is being ascertained by other consultants, one can state an aerated lagoon - irrigation system is a feasible means of treating either domestic sewage or mixtures of sewage and industrial wastewaters. This can be based on experimental and engineering data developed during the Muskegon County, Michigan Wastewater Treatment - Irrigation System Demonstration Study. The Merrimack River Basin, however, differs from Muskegon County in several areas relating to industrial wastewater loadings, composition and industrial load distribution. It is in these areas where an attempt will be made to extrapolate existing data and arrive at a reasonable estimate of the effects of industrial on any proposed LDS.

BKGD: In the Whitehall-Montaque area of the Muskegon County System, a separate, smaller facility, with chemical pre-treatment, was designed to handle a specialized situation. This approach will be considered for areas of the Merrimack River Basin [MRB].

SOURCES: a.) Winnepesaukee Area - The total industrial waste source in this area is from a tannery and an asbestos fiber plant. Depending on type of hides and tanning processes employed, wastewater could have numerous potential pollutants. Aside from the obvious chrome problem; salts, sulphides (mercaptans), fungicides, dyes, sulphates, acid, high B.O.D. and solids loadings could all be potential sources of trouble.

In the Whitehall-Montaque Study, the tannery wastewater was handled quite successfully with chemical pre-treatment. With the amount of dilution water in this basin, any pre-treatment decision would depend on accurate wastewater analyses.

There is little information available on what is involved at the asbestos processing plant. If mining is conducted, problems would differ from processing only.

b.) Lowell-Lawrence-Haverill Area - The industrial sources in this area are quite variable. In addition to normal industrial pollutants, the low water usage indicates high water recycle and potential problems from cooling tower blow-down [high salinity].

Among the thirty-three (33) industries in the area, twenty-seven (27) have flow data available. These industries give an average (1970) flow of 13.7 MGD. This can be compared to a municipal flow

SOURCES: of about 70 MGD (1970). The six industries without flow data probably do not contribute in excess of 3.5 MGD (about 25% of measured flow) based on limited information on type of industry. This could provide a 4:1 dilution of the industrial waste load by municipal sewage. At this dilution, pre-treatment may still be required, this decision again depending on LDS and where one prefers to manage industrial waste.

c.) Boston Area - Aside from direct discharges to Boston Harbor, and to several river basins, most industrial waste is combined with domestic sewage and handled by the two Boston Sewage Treatment Plants (primary). No data has been received on the concentration of pollutants, except B.O.D. and Suspended Solids, in the Boston primary effluent. Limited data from the Metropolitan Sanitary District of Greater Chicago on metal content of both preliminary influent and effluent does not indicate problems from toxic metals. The sludge concentrates several metals,but actual use of sludge as fertilizer has not caused any problems.

It should not be inferred that both Boston and Chicago have the same type sewage, but in absence of other information, it allows one to make an educated guess.

The direct discharges to the Harbor and adjoining river basins, from the nature of the industry, could represent high flows and potential problems in treatability.

TYPES OF INDUSTRIAL WASTE: From the industries listed in the Lowell-Lawrence-Haverill Area, Boston Harbor and those assumed to be in the Boston Metropolitan area, a number of industrial classifications can be reviewed for potential pollutants. These include the Dairy, Texile, Food Processing, Hat Manufacturer, Metal Plating and Processing,

TYPES OF INDUSTRIAL WASTE: Rendering, Tannery and Wool Scouring. The potential pollutants are covered quite adequately in the Corps of Engineers Permit Document.

The only addition might be pesticide residuals found in Wool Scouring effluent.

EFFECTS: Industrial Wastes can effect a LDS in any phase from collection and conveyance to irrigation and final discharge. The effects will vary from corrosion, to toxicity, to reducing effluent quality. Other problems can occur when toxic materials concentrate in the sludges. Which in turn hinders or limits means of ultimate waste disposal. It is really not necessary to list and evaluate every possible effect from every potential pollutant. Rather in the overall design of a system, means must be incorporated to control or eliminate adverse effects from known pollutants.

CONTROL: It is possible to control pollutants at any stage of a LDS. The deciding factor would depend on careful evaluation of LDS design, pollutant type and source, overall waste management philosophy and of course cost evaluation of alternatives.

A.) Winnepesaukee Area - Based on management philosophy, which I favor, maximum control of entire waste system by operating agency, it would appear pre-treatment on site at LDS would be preferred method of handling combined industrial and sanitary waste. When further is known about composition of various wastewater discharges, it may be possible to eliminate chemical pre-treatment and directly aerate influent.

CONTROL: B.) Lowell-Lawrence-Haverill Area

1.) Lowell has an industrial flow of about 1.7 MGD and a municipal flow of about 15 MGD. It seems from the types of industry (textile, food, tannery) and dilution from municipal flow, a LDS is feasible and probable without pre-treatment.

2.) Lawrence has a municipal flow of 32 MGD and an industrial flow of about 7 MGD. Except for the wool scouring and textile industry, the situation could resemble Muskegon County. The actual water usage per ton of product is lower; the solids, B.O.D., and dissolved solids are higher. This situation could indicate either pre-treatment at source or on-site.

The high use of water recycle in the area may cause problems in LDS, due to eventual salt build-up in soils.

3.) Haverill - The municipal flow is 21.6 MGD, and the industrial flow is incomplete but estimated at about 5 MGD. The industrial processes includes tanning, wool, scouring, paper and textile manufacturing.

This area is quite complex to evaluate, but may require pre-treatment at each plant or research to develop a common pre-treatment for on-site waste processing.

C.) Boston, Boston Harbor and adjoining River Basins - This is the most complex area and less data is available than for any of the other regions. In an area of this size, a major portion of industrial waste control must be accomplished at the source. An Industrial Waste Control Ordinance which can control pollutants to levels the system can accept without pre-treatment on-site is required. This does not mean one must adopt an existing ordinance which has been passed in another area, but rather one designed to

CONTROL: supplement the LDS. The LDS has many built-in safe-guards against shock loading and slugs of toxic materials, and the over-all removal of metals and solids gives greater freedom in developing a pollution control code which could simplify plant pre-treatment facilities. Any reductions in plant pre-treatment cost promotes the over-all acceptability of the plan to industry.

CONCLUSIONS: There is probably no situation where industrial wastes should be the limiting criteria or interfere with the over-all concept of a land disposal system for wastewater treatment. Any waste so difficult to treat, so as to cause excessive problems in operation or design, can always be controlled by proper in-house pre-treatment. As a general policy, I would prefer the operating agency to maintain control over the entire waste management system in its area.

Heavy metal removal can be assured by the LDS, either in the storage lagoon or in the soil system. Leaching of metals can be controlable by pH control in irrigation area or sludge disposal site.

Due to the long detention periods and combination of aerobic and anearobic treatment most materials are eventually degraded. Those materials which would concentrate in biological systems, should be removed between the various stages of the system.

Probably the worst problem in the Merrimack River Basin is not ordinary industrial wastes, but high dissolved solids from industrial sources. Little data has been presented to support

CONCLUSIONS: this conclusion, but even in the water-abundant Muskegon area, potential problems with sodium, calcium, chloride and sulphate ions where observed. Many industrial sources had wastes extremely high in dissolved solids. Limitations were placed on the number pounds of chloride which could be discharged per day.

While one could worry about pounds of pesticide [from wool scouring] or pounds of chromium [from tanning], the most serious problem I presently vision is dissolved solids build-up, especially when additional recycle is considered.

It is possible storm run-off could contribute sufficient low solids dilution water to allow a LDS to operate, but again additional data will have to be collected.

W. J. BAUER
CONSULTING ENGINEER - ENGINEERING HYDRAULICS
20 NORTH WACKER DRIVE
CHICAGO 6, ILLINOIS

May 10, 1971

Dr. John R. Sheaffer
Program Planning Group
ENGCW - S.A.
Rm. 4G-026 Forrestal Bldg.
Washington, D.C. 20310

Subject: Merrimack River Basin
Boston Metropolitan Area
Wastewater Management Study

Dear Dr. Sheaffer:

The purpose of this letter report is to present initial appraisals for land disposal wastewater management opportunities for the Merrimack River Basin -- Boston Metropolitan Area, by Wayne A. Cowlishaw and myself.

There are three forms which land disposal could take:

1) Land disposal of sludges produced by various types of treatment plants.

2) Land treatment and disposal of treated wastewaters in flooding basins.

3) Land irrigation of forest and crop lands by treated wastewaters.

In actual practice, a combination of all three forms will undoubtedly be utilized if ocean disposal is no longer permitted.

The schematic watershed map of Figure 1 shows the relative locations of the major features involved. A great deal of wastewater transportation is required because of the lack of adequate sites near

the urban centers. The design average rate of flow will be assumed to be 800 MGD for purposes of illustration. An additional increment for storm water mixed with the wastewaters is taken at 200 MGD, with a total average flow of 1000 MGD. This rate of storm water can be achieved through flow regulation by means of the proposed Boston Deep Tunnel Plan, which has been prepared by Camp, Dresser and McKee. The storage in the tunnels will permit control of the combined wastewater and storm flows to about 1.25 times the average dry weather flow, or (1.25) (800) = 1000 MGD.

The schematic illustration of Figure 2 shows how the soil will act upon wastewater or sludge applied to it. A major difference concerning the type of land treatment used lies in the quantity of horizontal water transportation needed, with the flooding basin system requiring the most, and sludge disposal requiring the least. For any of the land disposal forms discussed herein, the following soils and geological information is necessary:

1) The ability of the soils and underlying geological formations to move the water horizontally, sometimes called transmissibility, or transmissivity.

2) The ability of the soils and underlying geological formations to react chemically and physically with the materials carried in the wastewater.

3) The biological characteristics of the soil (plants and micro-organisms) which affect the conversions of wastewater into organic tissue.

A preliminary review of the types of wastewaters generated in the study area and their characteristics indicates that certain industries may be required to provide pre-treatment of their wastes prior to acceptance by the wastewater management systems. This could be accomplished in conjunction with "sewer use ordinances", and would be particularly applicable for smaller wastewater management systems that could not readily accommodate high BOD, or suspended solids loadings associated with some industrial wastes. Potentially toxic wastewater constituents would also need to be regulated by the "sewer use ordinances."

Treatment of the wastewaters through the use of lagoon facilities could be accomplished in areas located within the land disposal sites so as to provide for isolation of treatment facilities from the urban areas. Typically, the wastewaters would be conveyed, after any necessary pre-treatment, to the site for treatment in aerated lagoons followed by stabilization in storage lagoons, with the solids being deposited in the storage lagoons. Additional features at the site could include a settling lagoon to permit bypassing of the storage lagoons, and an outlet lagoon section of the storage lagoons to enable additional settling of the treated wastewaters prior to irrigation. Before irrigation, the wastewaters would be disinfected through chlorination.

There may be an alternative to the handling of the wastes by aerated lagoons in remote, large land disposal areas, particularly where the energy source for the aeration units may be restricted by accessibility or economics. Where adequate isolation is available, consideration could be given to the use of anaerobic lagoons, which can provide a high level of treatment at minimum operating costs. Since there is a potential odor problem with anaerobic lagoons, care would need to be exercised in considering their use. Plastic foam covers have been successfully used in assisting in temperature and odor control for anaerobic lagoons.

A typical flow diagram of lagoon treatment processes that could be employed is given in Figure 3.

We are looking forward to reviewing this information with you and other members of the Land Disposal Group at Amherst during the period beginning May 17.

Sincerely,

W. J. Bauer

W. J. Bauer

WJB:fb

Enc: 3 Figures

cc: W.A.Cowlishaw

WASTEWATER MANAGEMENT PROGRAM
MERRIMACK RIVER BASIN
BOSTON METROPOLITAN AREA

LAND DISPOSAL ALTERNATIVES

May 1971

Wayne A. Cowlishaw

Concept and Theory

Applying wastewater to the land for soil treatment requires that the wastes be treated prior to application. The prior treatment facilities needed are those that will produce a disinfected, "secondary effluent" quality of wastewater. More specifically, the wastewater should have a BOD_5 of 15 to 30 mg/l and total suspended solids of 15 to 30 mg/l. Based on 1968 information, only Maine of the three states involved in the study has a policy concerning the use of wastewater for irrigation.[(1)] Extensive work on wastewater irrigation and land disposal has been done in California, and this has resulted in the establishment of California standards for the use of wastewater for irrigation.[(2)]

---

[1] Coerver, James F., "Health Regulations Concerning Sewage Effluent for Irrigation," Municipal Sewage Effluent For Irrigation, Proceedings of Symposium held at Louisiana Polytechnic Institute, July 30, 1968.

[2] California Administrative Code, Title 17 - Public Health, "Statewide Standards for the Safe Direct Use of Reclaimed Waste Water for Irrigation and Recreational Impoundments," May 1968.

Other constituents, such as heavy metals, may also need to be controlled if their concentrations would not be acceptable based on public health or agricultural criteria. The Federal Water Pollution Control Administration has published a report which contains information regarding the quality of water used in irrigation.[1]

The use of wastewater lagoon treatment facilities to provide for the needed processing of the wastewaters prior to their use of irrigation involves two general types of lagoons. The two types are: (1) aerated lagoons that rely on mechanical equipment for aeration, and (2) storage or holding lagoons that function as waste stabilization ponds that rely on natural surface aeration and algae for aeration. A typical flow chart for a lagoon treatment system is shown on Figure 1.

Aerated Lagoons

Prior to conveyance to the aerated lagoons, the wastewaters would be screened to remove floating materials and other items that would be potentially harmful to mechanical equipment. The aerated lagoons are usually constructed with two or more separate cells. This permits passage of the entire wastewater flow through each of the cells in a series sequence, or the wastewater flow can be divided with portions going through one cell only as in a parallel operation. This

---

[1] Federal Water Pollution Control Administration, Water Quality Criteria, April 1, 1968.

PROJECT No. _____ INDEX PAGE _____
PROJECT NAME MERRIMACK RIVER - BOSTON
SUBJECT WASTEWATER MANAGEMENT STUDY

BY _____ DATE _____ CH'KD _____ DATE _____
SHEET _____ OF _____

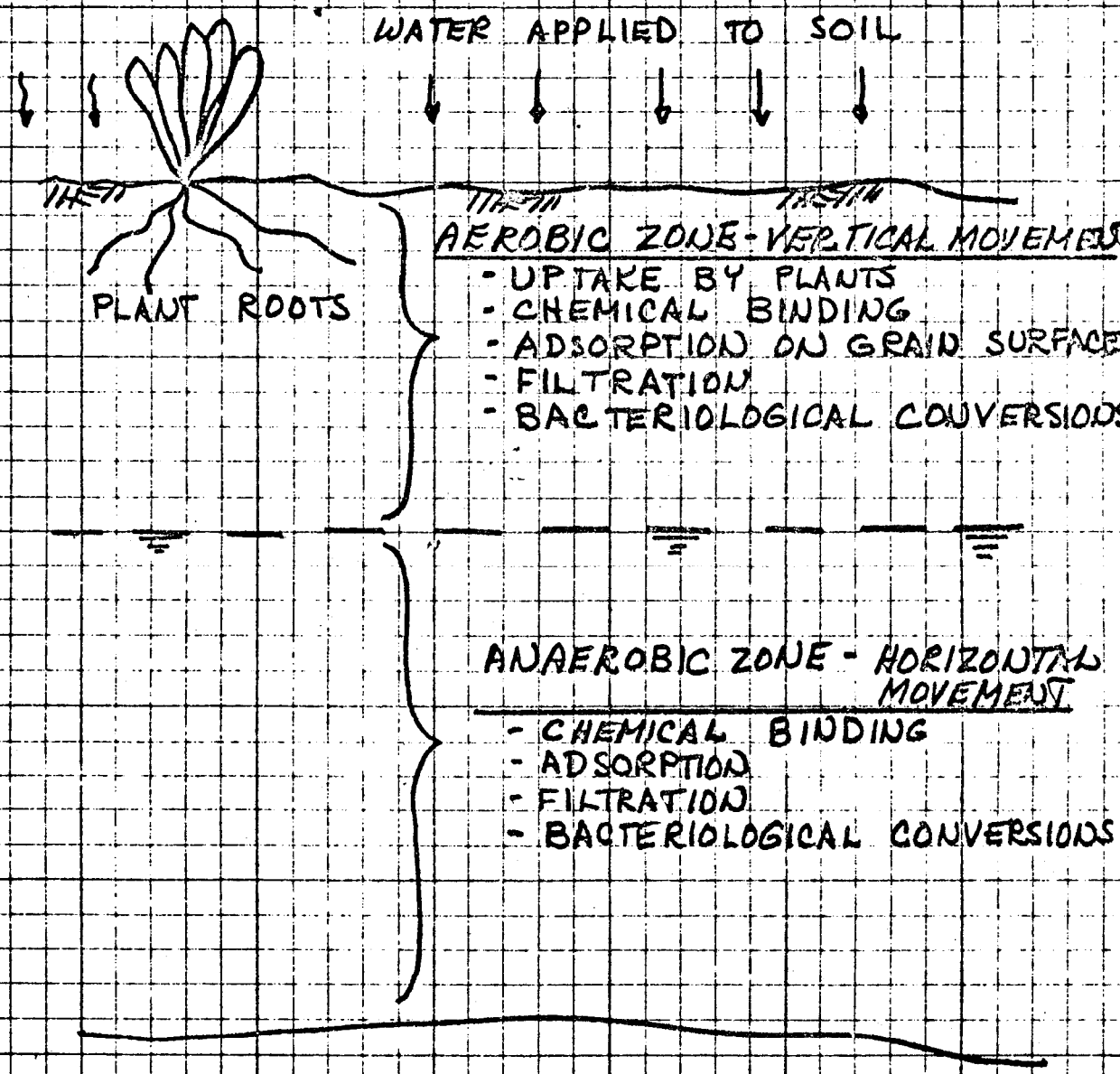
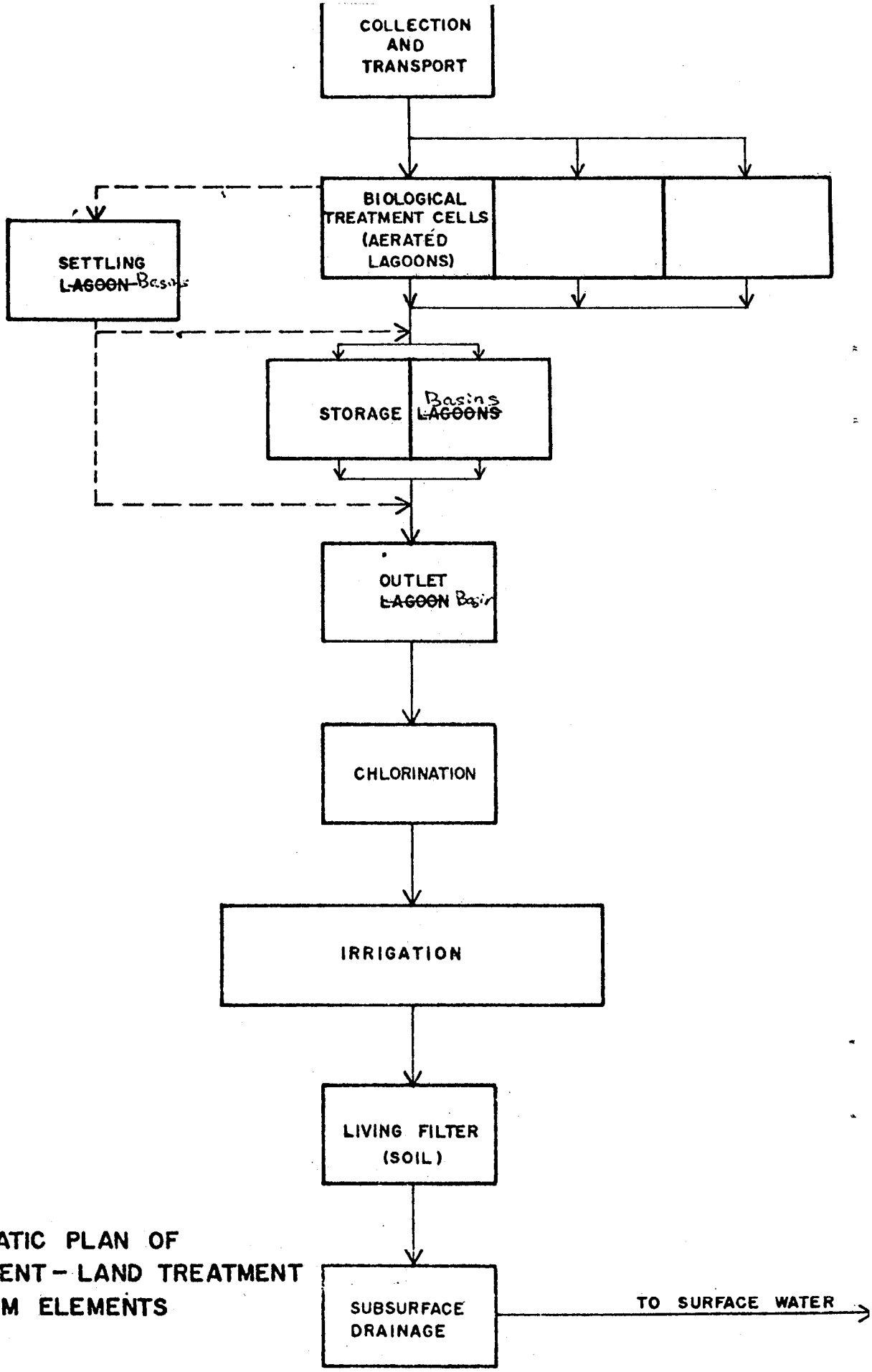


FIGURE 2

MECHANISMS INVOLVED IN LAND DISPOSAL



SCHEMATIC PLAN OF
L DON TREATMENT - LAND TREATMENT
SYSTEM ELEMENTS

feature permits flexibility of operation to obtain optimum treatment efficiencies as well as to provide for a method of containment of toxic waste spills that enter the system.

The main function of the aerated lagoons is to reduce the BOD of the wastewater to a level where the wastewater can be transferred to the storage lagoons or conveyed through the settling lagoon and then to the irrigation system. The reduction in BOD is accomplished by the addition of oxygen to the wastewater in the aerated lagoon by mechanical surface aerators either floating or fixed. Another important consideration in the design of the aerated lagoons is the management of the solids and sludge produced. Two approaches to this are to provide adequate mixing facilities to keep the solids in suspension or to permit deposition of the solids that are not maintained in suspension by the aerating equipment that is installed. If the solids are to remain in suspension, additional facilities such as mixers would be required. If the solids are permitted to settle, provisions for periodic removal would need to be provided.

Typical general bases of design for aerated lagoons are:

1. Two or more cells permitting series or parallel operation.

2. Depth -- 15 feet.

3. Detention time -- 3 to 4 days.

4. Aeration horsepower requirements -- 3 HP/100 lbs. BOD_5.

5. Mixing horsepower requirements -- 7 HP/MG.

6. Top of dike width -- 20 feet.

7. Dike side slopes dependent on soils used.

8. Freeboard -- 3 to 5 feet.

9. Interior side slopes provided with wave protection.

10. Bottom and interior side slopes provided with sealing materials.

Storage Lagoons

After treatment in the aerated lagoons, the wastewater can be conveyed to the storage lagoons or the settling lagoon. The main purpose of the storage lagoons is to provide containment for wastewaters during periods when irrigation cannot be accomplished, principally during the winter months. As an incidental part of the operation the storage lagoons can also provide further treatment of the wastes as well as a means of solids management. This aspect requires that the storage lagoons be designed to accommodate satisfactorily the BOD loading placed upon it. The area needed permits extensive distribution of the solids so that their depths of accumulation can be minimized. Collection and additional distribution of the solids can be achieved by use of a floating suction dredge.

Typical general bases of design for storage lagoons are:

1. Two or more cells.

2. Minimum depth -- 5 feet.

3. Minimum solids storage depth -- 2 feet.

4. Detention time -- 150 days.

5. Top of dike width -- 15 feet.

6. Dike side slopes dependent on soils used.

7. Freeboard -- 3 to 5 feet.

8. Maximum BOD_5 loading -- 15 to 20 lbs./acre/day.

9. Interior side slopes provided with wave protection.

10. Bottom and interior side slopes provided with sealing materials or groundwater management system provided.

Where depths greater than 10 to 15 feet are to be used, additional considerations will need to be given to maintaining adequate dissolved oxygen at various levels. This may mean that a higher level of treatment of the wastes prior to conveyance to the storage lagoons may be required or that supplemental aeration facilities would be needed in the storage lagoons.

Settling Lagoon

The settling lagoon permits the solids to be settled and concentrated in a smaller area than that of the storage lagoon. After passing through the settling lagoon, the wastewater can then be conveyed to the storage lagoons or the outlet lagoon. Typically, the settling lagoon would have a detention time of 12 to 24 hours. A suction dredge would be needed to collect the solids and convey them to the storage lagoon or other facility for additional treatment and stabilization. The design of the settling lagoon would be comparable to that for an aerated lagoon cell.

Outlet Lagoon

Prior to passage to the disinfection facilities, the treated wastewater is passed through an outlet lagoon. The function of this facility is to enable withdrawal of water isolated from the main storage lagoons for irrigation. This permits additional operating flexibility by providing for a certain level of irrigation water quality control. An additional feature of this lagoon is that it will allow supplemental deposition of solids prior to withdrawal. Typically, the outlet lagoon would have the general design features of the storage lagoons, but with a detention time of 3 to 4 days.

Disinfection

Before distribution to the soil treatment system, the wastewater would be disinfected to eliminate possible public health problems in managing the wastewater on the site. The measure of the effectiveness of this process is usually related to the remaining total coliform or fecal coliform organisms in the water after disinfection. Chlorination facilities are generally provided for this purpose and include equipment to permit applying a dosage rate up to about 15 mg/l to the wastewater, with a normal operating range expected to be between 2 to 5 mg/l. Also a part of the facilities would be a chlorine contact tank or channel which would have a minimum chlorine contact time of 15 minutes at the peak irrigation rate. After disinfection, a period of time for de-chlorination of the wastewater would be provided by the irrigation distribution system so that there would be no free residual chlorine in the water to interfere with the soils and agricultural biological processes.

Soils Treatment Types

There are three basic types of soils or land treatment of wastewater. They are: (1) agricultural-irrigation, (2) infiltration-percolation, and (3) spray-runoff.

The agricultural-irrigation system involves the use of piping and machines that permit the spraying of wastewater at the rate of about 8.5 feet per year over an area where crops are grown. The biological, chemical, and physical actions of the soils and plants provides the treatment of the wastes. Extensive work on this type of treatment has been done at Pennsylvania State University and Muskegon County, Michigan.[1] In the infiltration-percolation method,

---

[1] Parizek, R. R.; Kardos, L. T.; Sopper, W. E.; Myers, E. A.; Davis, D. E.; Farrell, M. A.; and Nesbitt, J. B., Penn State Studies Wastewater Renovation and Conservation, Pennsylvania State University Studies No. 23, University Park, Pa.

Bauer Engineering, Inc., Muskegon County Plan for Managing Waste Water, May 1969.

Bauer Engineering, Inc., Appendix C - Design Basis for Muskegon County Plan for Managing Waste Water.

Muskegon County Board and Department of Public Works, Engineering Feasibility Demonstration Study for Muskegon County, Michigan, Wastewater Treatment-Irrigation System, September 1970, Federal Water Quality Administration Research and Demonstration Project No. 11010 FMY.

a high rate of wastewater, about 150 feet per year, is applied to flooding basins where treatment is achieved by the soils. The basins are inundated for about two weeks and then permitted to dry for a one to two week period as in the Phoenix, Arizona Salt Creek (Flushing Meadows) Project.[1] The spray-runoff system is used where the soils are relatively impermeable as in the Campbell Soup Company project in Paris, Texas.[2] The rate of application is comparable to that for the agricultural-irrigation system, but instead of percolating through the soil, the wastewater flows horizontally through the surficial soils and plant roots where the biological, chemical, and physical reactions occur.

---

[1] Bouwer, Herman, "Ground Water Recharge Design for Renovating Waste Water," Journal of the Sanitary Engineering Division, ASCE, SA1, February 1970, pp. 59-74.

[2] Thomas Richard E.; Law, James P., Jr.; and Harlin, Curtis C., Jr., "Hydrology of Spray-Runoff Wastewater Treatment," Journal of the Irrigation and Drainage Division, ASCE, IR 3, September 1970, pp. 289-298.

Law, James P., Jr.; Thomas, R. E.; and Myers, Leon H., Nutrient Removal from Cannery Wastes by Spray Irrigation of Grassland, November 1969, Federal Water Pollution Control Administration Research Project No. 16080.

Applicability

The use of lagoons for treatment and storage of wastewaters prior to land treatment in the Merrimack River-Boston area is feasible. The constraints on the design of such facilities would be principally related to winter operation and the large volume of storage that would be required for the Boston Metropolitan area. In this case, it may be preferable to consider the construction of a reservoir or reservoirs having depths up to 50 to 100 feet to reduce the amount of area required and to take advantage of the natural topography. If this type of storage is provided, a higher level of treatment of the wastewaters would probably be needed. This could be accomplished by the separate removal and treatment of solids so that their oxygen demands in the storage reservoirs would be eliminated. An additional feature that could be incorporated in the storage reservoirs is diffused air piping to enhance oxygen transfer throughout the depths as well as to help minimize ice cover. This will permit better control of the water quality in the reservoirs to alleviate possible odor problems during the spring "ice break-up" period.

Alternatives to the provision of large storage facilities exist in the use of land treatment during cold weather periods. The amount of storage volume required could be significantly reduced by using the infiltration-percolation method of land disposal during times when agricultural-irrigation or spray-runoff land treatment are not appropriate.

Design Criteria

Included in Table I are the basic annual average design flows for the areas considered in the Merrimack River Basin and Boston Metropolitan Area. Since most of the municipalities have combined sewer systems, an allowance for stormwater has been included based on the assumption that storage facilities will be provided to receive combined sewer overflows, thereby permitting regulation of the stormwater input to the collection and treatment system. Table II lists the estimated future BOD and Suspended Solids loadings from municipal and industrial sources for the Lake Winnipesaukee, Lowell-Lawrence-Haverhill, and Boston areas.

Using the general basis of design given in the "Concept and Theory" section of this report, the lagoon treatment facilities needed for the Lake Winnipesaukee, Lowell-Lawrence-Haverhill, and Boston areas are summarized in Table III.

Effluent Quality

Since there will be a variability in treatment performance within the lagoon system depending on a variety of factors such as temperature and detention time, the parameters and constituents of the treated wastewater withdrawn for soils treatment will have a range of values. Because the quality requirements for wastewater to be used in land disposal can be generally classed as those for a disinfected "secondary effluent," this variability in performance will not adversely affect the system, unlike that for certain unit processes where stricter influent

TABLE I

ESTIMATED ANNUAL AVERAGE FLOWS
FOR 1990 AND 2020

Area	Annual Average Flows - MGD							
	Municipal		Industrial		Stormwater[1]		Total	
	1990	2020	1990	2020	1990	2020	1990	2020
Lake Winnipesaukee	7	14	3	1	2	3	12	18
Lowell-Lawrence-Haverhill	105	146	16	17	26	36	147	199
Boston	500	710	(2)	(2)	125	178	625	888
Concord-Manchester	52	94	5	7	13	24	70	125
Fitchburg	17	27	18	23	4	6	39	56
Nashua	30	52	5	8	8	13	43	73
TOTALS	711	1043	47	56	178	260	936	1359

TABLE II

ESTIMATED BOD AND SUSPENDED SOLIDS
LOADINGS FOR 1990 AND 2020
(MUNICIPAL AND INDUSTRIAL ONLY)

AREA	BOD				Suspended Solids			
	1990		2020		1990		2020	
	Lbs./Day	mg/l	Lbs./Day	mg/l	Lbs./Day	mg/l	Lbs./Day	mg/l
Lake Winnipesaukee	13,433	161	16,395	131	12,783	153	15,442	124
Lowell-Lawrence-Haverhill	218,000	216	337,000	248	247,000	245	330,000	243
Boston	880,000	29	1,225,000	205	1,370,000	330	1,690,000	285

TABLE III

APPROXIMATE DESIGN FEATURES FOR
LAGOON TREATMENT AND STORAGE FACILITIES
1990 FLOWS

Item	Lake Winnipesaukee	Lowell-Lawrence Haverhill	Boston
Aerated Lagoons			
Volume - MG[1]	36	441	1,875
Surface Area - Acres	8	90	385
No. of Cells	3	9	21
Total Aerator H.P.	400	6,540	26,400
Total Mixer H.P.	250	3,090	13,125
Settling Lagoons			
Volume - MG[2]	12	147	625
Surface Area - Acres	2.5	30	130
No. of Cells	1	3	6
Storage Lagoons			
Volume - MG	1,500	18,150	75,000
BOD Loading - Lb./Day[3]	3,360	54,500	220,000
Surface Area - Acres[4]	190	3,115	12,570
No. of Cells	2	5	15
Outlet Lagoons			
Volume - MG[5]	36	441	1,875
Surface Area - Acres	8	90	385
No. of Cells	1	5	20
Chlorination Facilities			
Max. Feed Rate - Lbs./Day[6]	1,670	20,850	88,400
Chlorine Contact Volume - MG	210,000	2,610,000	11,040,000

(1) 3-Days Detention

(2) 24-Hour Detention

(3) Assuming 75% reduction in influent BOD

(4) Based on loading of 17.5 lbs. BOD_5/acre/day

(5) 3-Days Detention

(6) Based on 10 mg/l.

quality control is required. A listing of parameters and constituents with their expected values for the various management systems is given below:

Parameter or Constituent	Value mg/l.
BOD_5	10 - 20
SS	10 - 20
COD	20 - 40
N	10 - 20
P	2 - 10

Perhaps the two key factors concerning the operation of the lagoon treatment systems that are of more significance than the effluent quality per se are the maintenance of a relatively odor-free plant and the nutrient values of the wastewater. If the lagoon systems as previously described are operating without nuisance conditions, then the quality of effluent can generally assume to be satisfactory. The nutrients and their form within the effluent (nitrogen, phosphorus, potassium, etc.) will usually be the most important consideration regarding the agricultural irrigation type of soil treatment system.

Experience on Processes

Aerated Lagoons

The first aerated lagoon experiments were conducted in 1957, and the initial installations came on line in the 1960's. During the period of development in the 1960's floating mechanical surface aerators were perfected for use as the main oxygen transfer method in the aerated lagoons. With this experience has evolved various bases of design for these treatment cells. Where industrial wastes are included in the wastewaters, laboratory or pilot-plant operations are often tested to demonstrate the treatability of the wastes and to help select the optimum design.

Extensive use has been made of aerated lagoons for treatment of domestic and industrial wastes. A number of papers concerning their utilization were presented at the Second International Symposium for Waste Treatment Lagoons held in Kansas City, Missouri on June 23-25, 1970. This conference was sponsored by the Missouri Basin Engineering Health Council and the Federal Water Quality Administration, and the papers presented at the conference were published. Of particular interest in the conference proceedings is a paper reviewing the current status of aerated lagoons.[1] In this article and other articles in the proceedings, the development of aerated lagoons is discussed and various existing installations are reviewed.

---

[1] McKinney, Ross E., "State of the Art-Aerated Lagoons," Second International Symposium for Waste Treatment Lagoons, Kansas City, Missouri, June 23-25, 1970, Sponsored by Missouri Basin Health Council and Federal Water Quality Administration.

Problems in the operation of aerated lagoons in northern climates have been chiefly attributable to the following factors:

1. Cold weather problems including icing of aerators.

2. Inadequate solids or sludge management provisions

Cold weather problems can be managed by proper design and recognition of the elements involved. Required are the installation of aerators designed to minimize icing, the operation of the cells to reduce heat loss (such as reducing the number of cells used in the winter because of attenuated oxygen demands) and consideration of adding heat to the raw wastewater by exchange with a waste thermal source (such as a power plant). Solids problems, such as those that have occured at the Packaging Corporation of America's Rittman, Ohio plant, can be corrected by providing adequate mixing in the aerated lagoons to maintain the solids in suspension and minimize deposition. The key is to recognize the quantities of solids and sludges that will be conveyed to the lagoons and to provide an adequate management program, through design and operation, for their treatment and disposal.

Storage Lagoons

As previously mentioned, the storage lagoons will function in many respects as waste stabilization or oxidation ponds. There are presently about 3,700 ponds that have been constructed within the United States since the 1940's including a great number in northern climates. Experience in their operation has led to the development of design criteria to permit satisfactory operation and results. Extensive information regarding their use is contained in papers presented at the

Second International Symposium for Waste Treatment Lagoons held in Kansas City, Missouri on June 23-25, 1970. Of particular interest in the conference proceedings is a paper covering the status of oxidation ponds.[1]

Operational problems concerning waste stabilization ponds have evolved around nuisance conditions associated with overloading or inadequate design and the "Spring ice break-up." Odors can be controlled by proper design and loading for the climate prevailing in the area where the lagoons are to be constructed. Nuisance conditions connected with the "Spring ice break-up" period can be minimized or eliminated by the proper management of solids, a careful review of the BOD loading, or the installation of diffused air piping or other mechanical means to reduce the amount of ice cover and maintain dissolved oxygen at various levels.

Lagoons - General

The other components of the lagoon treatment system such as chlorination facilities have been demonstrated to perform satisfactorily when properly designed and operated. Additional references on lagoon treatment are included in Pages A-19 through A-28 of Appendix A.

---

[1] Vennes, John W., "State of the Art-Oxidation Ponds," Second International Symposium for Waste Treatment Lagoons, Kansas City, Missouri, June 23-25, 1970, Sponsored by Missouri Basin Health Council and Federal Water Quality Administration.

Soils Treatment Systems

Over 400 municipalities in the United States utilize some form of soils treatment or land disposal of wastewaters. Included on Pages A-1 through A-14 of Appendix A is a list of references concerning land disposal. Three other publications also contain articles and bibliographies on soils treatment of wastewater[1]

The Pacific Southwest Inter-Agency Committee report contains an extensive listing of soils treatment facilities, and this listing is included as Appendix B.

Problems in operations of soils treatment systems have been principally related to design factors. These include:

1. Insufficient soils permeability.
2. Inadequate subdrainage management and facilities.
3. Inadequately treated wastewaters.
4. Overloading of the soil system.

---

[1] Wilson, C. W. and Beckett, F. E. (Editors), Municipal Sewage Effluent for Irrigation, Louisiana Polytechnic Institute, Symposium held on July 30, 1968.

McGauhey, P. H. and Krone, R. B., Soil Mantle as a Wastewater Treatment System, December 1967, Sanitary Engineering Research Laboratory, University of California, Berkeley, SERL Reprot No. 67-11.

Pacific Southwest Inter-Agency Committee, An Inventory of Wastewater Reclamation and Re-use in the Pacific Southwest Area, 1970.

Apparently, high groundwater problems in the Lubbock, Texas installation resulted from the high loadings on the soil and inadequate subdrainage facilities.

It is recognized that there is much to be learned about the complex physical, chemical, and biological reactions occuring in the soils system so as to be able to more readily predict its treatment characteristics. However, with the proper consideration of the presently available soils information and adequate design, the installation and operation of suitable soils treatment systems is entirely feasible.

Capital and Operating Costs

Costs for the construction and operation and maintenance of the lagoon treatment and agricultural-irrigation systems for Lake Winnipesaukee, Lowell-Lawrence-Haverhill, and Boston have been estimated and are presented for 1990 flows in Table IV. The construction costs do not include engineering, legal, land acquisition, sewers, force mains, or pumping stations. Also included in Table IV are the estimated total land requirements for agricultural-irrigation soil treatment facilities, including an allowance for lagoons.

Process Dependability

Lagoon Treatment

A sufficient amount of operating experience has demonstrated that the aerated and storage lagoons, when properly designed and

TABLE IV

APPROXIMATE LAGOON TREATMENT - AGRICULTURAL IRRIGATION SYSTEM
COSTS AND LAND AREAS

System	Estimated Construction Costs[1]			Estimated O&M Costs-Cents/1000 Gal.			Total Land Required
	Lagoons	Irrigation	Total	Lagoons	Irrigation[2]	Total	Acres[3]
Lake Winnipesaukee	$ 2,500,000	$ 3,700,000	$ 6,200,000	$0.105	$0.045	$0.15	2,500
Lowell-Lawrence-Haverhill	16,100,000	24,100,000	40,200,000	0.05	0.02	0.07	30,000
Boston Metropolitan Area	68,000,000	101,000,000	169,000,000	0.05	0.02	0.07	120,000

(1) - Does not include engineering, legal, land acquisition.
(2) - Does not include farming expenses or revenues.
(3) - Includes allowance for lagoons, roads and other non-irrigated areas.

operated, have overall adequate process dependability. Because of the relatively large volumes of water involved, long detention times and flexibility of operation, the lagoon processes are not as sensitive to time, pH or flow rate as other unit processes.

Perhaps the key factor in influencing lagoon process efficiency is temperature. It has been well documented that aerated lagoon BOD removals will be less during cold weather.[1] In storage lagoons, the result of extended cold weather can be the formation of an ice cover which will inhibit oxygen transfer to the stored water with the possible development of anaerobic conditions. The removal of the ice during "spring break-up" can cause the release of nuisance products of anaerobiosis.

Since the net result of cold weather operations is to increase the ultimate BOD loading on the storage lagoons and decrease the available dissolved oxygen therein, provisions can be made in the design and operation to compensate for this. Included can be: (1) operation of the aerated lagoons to conserve as much heat as possible, (2) the introduction of a waste thermal load into the system, (3) deferring solids transfer to the storage lagoon, and (4) the installation of diffused air piping in the storage lagoons.

Soil Treatment

Although soils systems can be overloaded and still be capable of removing certain wastewater constitutents, their proper design and

---

[1] Bartsch, Eric H. and Randall, Clifford W., "Aerated Lagoons - A Report on the State of the Art", Journal Water Pollution Control Federation, 43, No. 4, pp 699-708 (April 1971).

operation is important in maintaining process efficiencies. This is particularly true concerning the management of nitrogen, phosphorus and heavy metals. Depending on the type of soil treatment being used, different nitrogen removals will be encountered. It is also anticipated that during the operation of a given soils system there will be some variation in nitrogen removals. The important point to be made is that prior to development of a land disposal or soils treatment system, the nitrogen reduction requirements should be clearly defined so that the facilities can be adequately designed.

Physical factors influencing the dependibility such as soils clogging and groundwater mounding can be managed through soil renovation techniques and proper design and operation of a subdrainage system.

******

PROJECT No. _____ INDEX PAGE _____

PROJECT NAME MERRIMACK RIVER - BOSTON

SUBJECT WASTEWATER MANAGEMENT STUDY

BY _____ DATE _____ CH'KD _____ DATE _____

SHEET _____ OF _____

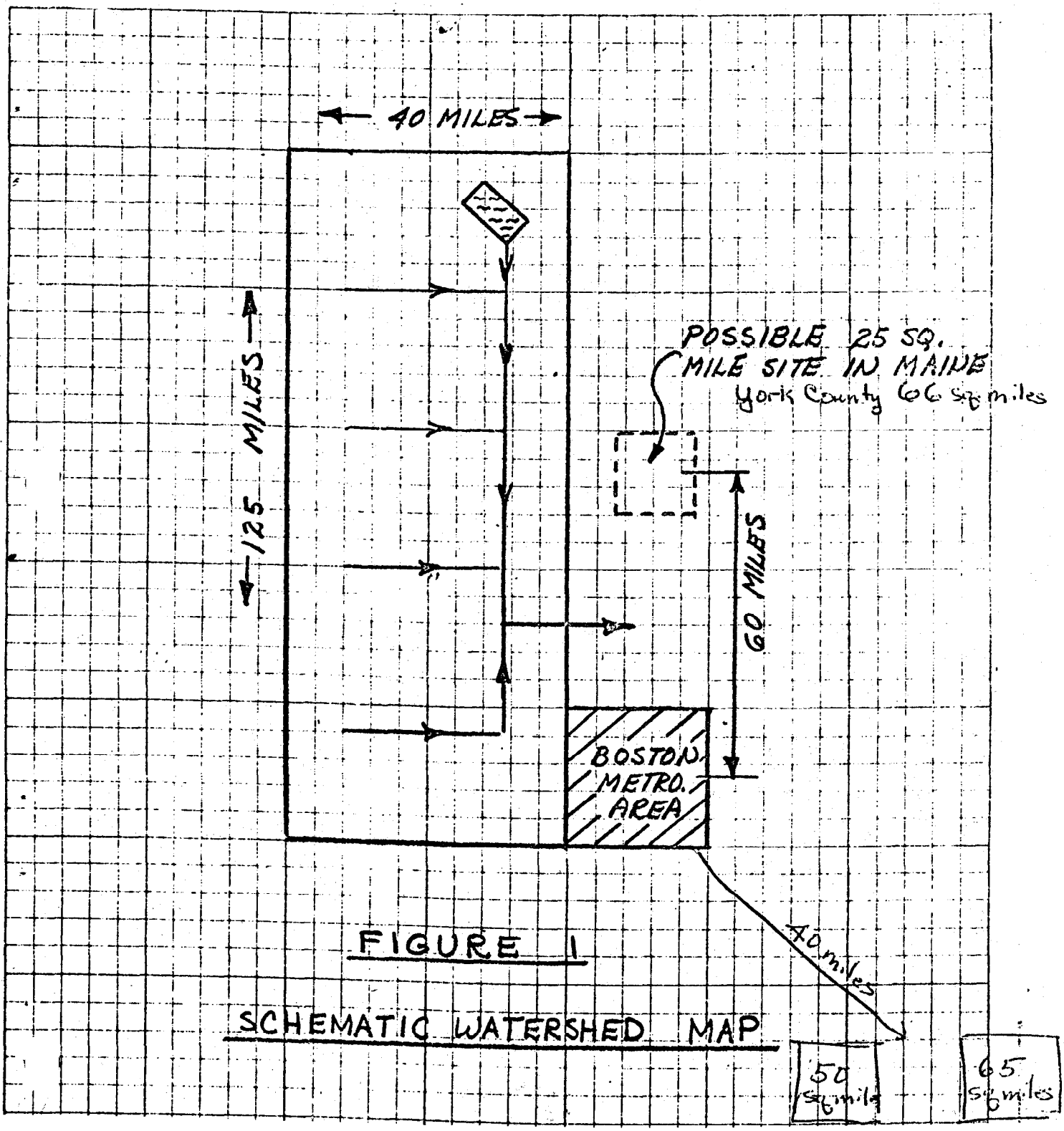


FIGURE 1

SCHEMATIC WATERSHED MAP

Project No. 1208.001

June 6, 1971

WASTEWATER MANAGEMENT PROGRAM
MERRIMAC RIVER BASIN
BOSTON METROPOLITAN AREA

STORAGE AND TRANSPORATION SYSTEMS

June 1971

W. J. Bauer

Water storage and transportation costs, urban areas of Merrimack River Basin and Boston area.

Given:

1. Estimated average annual flows of TABLE I of W.A.Cowlishaw paper LAND DISPOSAL ALTERNATIVES, May 1971.
2. Existing combined sewer systems throughout most of presently urbanized areas.
3. Desire to achieve a high degree of wastewater renovation to permit reuse of water for domestic and industrial use.
4. Desire to consider use of land treatment potentials such as described in the Cowlishaw paper.

Find:

1. Storage required to regulate flow variations produced by stormwater flows.
2. Combination of conveyance and pumping structures required to transport the wastewater to storage facilities.
3. Combination of conveyance and pumping structures required to transport the wastewater from the storage facilities to the treatment facilities.

Storage System

Background

Statistical studies of 20+ years of hourly rainfall records in Chicago were used as input to a computer model simulation study to discover the relationship between volume of storage and frequency of exceeding this volume of storage with various assumed rates of flow through treatment facilities. Similar studies should be made for the urban areas of Boston and other cities involved in this study. Such studies have been made in the studies of Camp, Dresser & McKee for the Deep Tunnel Plan for the Boston area.) In the absence of having time to assemble the data required to establish such relationships with precision, the following estimate is drawn from experience with the Chicago studies:

TABLE A

Item	Alternatives A	Alternatives B
Average flow of storm water spread over the year	157 MGD	157 MGD
Actual flow rate when sewers operate to capacity	25,000 MGD	25,000 MGD
Storage provided	0	8,800 MG
Flow rate allocated for treatment of storm flows	2,198 MGD	260 MGD
Flow rate allocated for treatment of nonstorm flows	1,099 MGD	1,099 MGD
Total flow rate allocated for treatment of combined flows	3,297 MGD	1,359 MGD

The benefit of using storage is very evident in this TABLE A. The incremental cost of the 1938 MGD of treatment capacity at $500,000 per MGD would be about $969 million. At $10 per cubic yard of rock

excavation, which includes provision for solids handling and aeration, the cost of 1 million gallons of space would be about $50,000. The cost of the conjectured 8800 MG of storage would be $440 million. The use of storage would also permit a higher degree of reliability of the system by providing a space for accidental "spills" which in the absence of storage are commonly discharged with little or no treatment directly into the receiving water body.

It will be assumed in the subsequent portions of this memorandum that the required storage -- on the order of 8800 MG for 1359 MGD of treatment capacity -- would be provided in such locations and in such forms as would prove ultimately to be the best solution to the total water resource management problem in the Boston vicinity.

Forms of Storage

Studies made in connection with the Deep Tunnel Plan for the Chicago area have indicated two practical forms for large volume storage:

1) Underground excavations in large mined-out galleries in self-supporting rock formations under the urbanized areas.
2) Surface storage in diked-off space is available. The first type can function without pumping at high rates. Because of this, the practical form of storage will almost certainly include a substantial quantity of this underground storage. If surface storage is to be practical, it should probably be associated with a companion underground storage and possibly a pumped-storage hydroelectric facility.

Solids Handling and Aeration

The maximum volume of storage amounts to approximately 8 days' flow from domestic and industrial sources. This volume would be passed through the treatment facilities over a 32-day period along with the normal day-to-day flow from nonstormwater sources. This maximum duration of storage would of course be associated with unusually rare, large storms. However, relatively common rains would require a storage of a week or so. It is apparent that provision must be made for aeration, sedimentation and removal of flotables in the storage facilities. These treatment facilities correspond to an extended primary treatment for that portion of the flow which would be stored. The underground storage facilities could take the form of chambers 50 feet in width and 100 feet in height of which a depth of 90 feet would be useful storage. The upper 10 feet could constitute an air passage for ventilation to provide a source of oxygen for surface aerators if these were to be used. Such chambers would store about 33,000 gallons per foot of length, so 30 feet would be required for each 1 MG of volume. A total length of 264,000 feet, or about 50 miles of such chambers would be required. Constructed in a grid pattern at 100' centers, there would be 50 miles of such chambers in one square mile of area. Consequently, 640 acres of area, divided into a number of strategic locations, would suffice for locations of these storage facilities. It is unlikely that there would be 2 or 3 underground storage areas for Boston, plus one each for Lowell, Lawrence, Haverhill, Concord, Manchester, Fitchburg and Nashua, a total of, say, 10 locations. As Boston constitutes about 2/3 of the flow, it is expected that each of the three Boston facilities would cover about 160 acres. This is seen to be about 1/2 mile square in area, which is not an unreasonable size for a mine. The 1/2 mile square area could be organized into one long central gallery with 1/4 mile long galleries tributary to both sides at 100' intervals. In each

1/4 mile long gallery there would be a scraper mechanism to move the deposited solids along to the central gallery from which they would be pumped to digesters either underground or on the surface. Flotables would be screened at points of entrance of main tunnels to the storage facility; screenings would be collected and comminuted and pumped to the same digesters.

The maximum aeration demands of the maximum volume of 8800 MG of stored wastewater would be on the order of 5 million pounds of oxygen per day. This may be compared to a typical daily demand on the order of about 1,500,000 pounds of oxygen for the treatment facilities with an average flow rate of 1356 MGD. The 5 million pound demand rate for oxygen would, however, be a relatively rare occurrence. Consideration should be given to the possibility of using stored tonnage oxygen as the source for this relatively rare, high-rate oxygen demand. (The reader is reminded that these figures correspond to the year 2020 condition of Cowlishaw's TABLE I, and are actually the sum of demands at a number of locations.)

This discussion is not intended to present design values for facilities, but merely to indicate the nature of the considerations which must be given in the final design of storage facilities for handling overflows from combined sewers. By locating these facilities underground as large mined-out areas in self-supporting rock, they can be located to minimize the cost of transporting the water to them and can be constructed with a minimum of disruption of existing densely occupied surface areas.

The attached copy of a paper on "The Role of Storage in Sewage Treatment Plant Design" is appended for the convenience of the reader who would like to read further on this subject.

Transportation System

Gravity Tunnels

As it is almost certain that underground storage in the rock would be used as the most economical method of solving the problem of overflows from combined sewers, it is logical to explore the possible type of tunnels in the rock which could be used for transportation of the wastewater. Two types of transportation are required.

1. High-rate transportation to convey the discharge from combined sewers to points of storage. These would be designed to receive the full rate of discharge of the combined sewers under storm conditions. It is estimated that the 1099 MGD average flow from nonstormwater sources comes from 165,000 acres of urbanized areas, and that the corresponding peak flow rate of all of the associated combined sewers would be about 40,000 cubic feet per second.

2. Low-rate transportation of the regulated flows out of the storage facilities to the place of treatment. This would occur at a rate of about 1359 MGD = 2040 cfs. This is seen to be very much smaller than the unregulated flow rate, again emphasizing the effect of storage.

It is obvious that the distance through which the high-rate transportation would be provided should be minimized. This can be accomplished by locating the storage facilities close to the outlets of present combined sewers. This would indicate the need for tunnels along the rivers into which present sewers discharge, with vertical shafts connecting the outlets of these sewers with the new intercepting tunnel. This new tunnel would function as both a conveyance

tunnel and as part of the required storage. By way of illustration, the larger storage chambers serving, say, 25% of the total area would receive inflows at the rate of, say, 10,000 cfs each. Assuming two main inflow tunnels to each such storage chamber, the capacity of each would be 5,000 cfs. Assuming a design velocity of 10 fps, the corresponding flow area would be 500 sq.ft. corresponding to a tunnel 25 feet in diameter. This is a workable size for efficient tunnel construction. The upper end of such a tunnel would of course be smaller. The hydraulic gradient required to produce 10 fps in a 25' diameter tunnel is about 5 feet per mile. Assuming such a tunnel would serve an area of 30 square miles, its length could be expected to be about 6 miles, and the corresponding friction loss about 30 feet. This is probably smaller than would be found economical, but serves to illustrate the magnitude of the head involved.

Two tunnels of this size would discharge into each of the main storage facilities, with screening, automatic raking, comminuting and pumping of comminuted flotables being provided at each point of entry.

From each such storage facility, the flow would be pumped at a peak rate of about 500 cfs. Assuming a design velocity of 5 feet per second, a tunnel of 11 feet in diameter is indicated. Assuming a distance of 50 miles to the treatment facility, the length diameter ratio is 260,000 ÷ 11 = 24,000. With a friction factor of 0.02 and velocity head of 0.4', the friction head loss would be 480 x 0.4 = 192'. The corresponding friction horsepower would be about 500 cfs x 192'/6 = 16,000 hp. Assuming a typical gravity lift of 200 feet, the total pumping horsepower would be 32,000 hp.

Based upon the foregoing conceptual analysis of the transportation system, one may construct the preliminary estimate of cost summarized in TABLE B. It can be noted that this same system may be used to receive, store, treat and transport wastewater to either a conventional or a land system of treatment. The only difference in cost would be in Items (5) and (6) which account for about 1/3 of the total cost. With a conventional treatment system a shorter force main may be required, reducing the length of Item (6) and the head of Item (5).

TABLE B

PRELIMINARY COST ESTIMATE OF STORAGE TRANSPORTATION FACILITIES

Item	Description	Quantity	Rate	Amount
(1)	One of 3 main storage facilities serving Boston	2200 MG	$50,000[c]	$110,000,000
(2)	Same as (1)	2200 MG	50,000[c]	110,000,000
(3)	Same as (1)	2200 MG	50,000[c]	110,000,000
(4)	Tunnels feeding (1), (2) and (3) at rates of 10,000 cfs to each, average diameter = 20 feet	36 mi.	$3,000,000	108,000,000
(5)	Pumping stations at each of (1), (2) and (3), averaging 500 cfs @ 400' head each.	100,000 hp.	$300	30,000,000
(6)[b]	Force mains averaging 50 miles in length x 11' diameter for each of (1), (2) and (3)	150 mi.	$1,500,000	225,000,000
(7)	Subtotal			693,000,000
(8)	Add 33% for smaller facilities of 7 smaller cities			231,000,000
(9)	Subtotal			$924,000,000[a]

Notes:

(a) Engineering, legal, financial, right-of-way and land costs are not included.

(b) Note Item (6) is about 1/3 of the total cost.

(c) Includes solids handling and aeration equipment.

Summary

The following rough cost estimates compare an agricultural irrigation system, plus the associated storage and transportation system with an advanced waste treatment system, plus a storage and transportation system:

TABLE C

1359 MGD Regulated Systems

	Agricultural Irrigation System	Advanced Waste Treatment System
Storage and Transportation	$924,000,000	$650,000,000
Treatment System	215,000,000	680,000,000
TOTALS	$1,139,000,000	$1,330,000,000

These figures merely indicate costs to be comparable. Factors other than costs should also be assessed in detailed feasibility studies of these alternatives.

The attached copies of previous reports on related subjects may be useful.

(1) The Role of Storage in Sewage Treatment Plant Design.

(2) Storm Drainage Aspects of the Deep Tunnel Plan (A.I.S.I. Paper)

Report

to

North Atlantic Division
Corps of Engineers
U.S. Army

on

WASTE TREATMENT BY REVERSE OSMOSIS

by

Frank Agardy, Ph.D.
URS Research Company
San Mateo, California

May, 1971

Reverse Osmosis

Section 1

STATEMENT OF PROBLEM

The basic purpose of this review is to develop a technically realistic wastewater management problem for the Merrimack River Basin. Specifically, consideration is to be given to three sub-regions within the Basin, namely, the Boston Metropolitan Area, the Lowell-Lawrence-Haverhill Area, and the Winnipesaukee Area. For each of these areas, a projected design is based on the 1990 wastewater flow information. Treatment design schemes consider both the handling of municipal wastes and the handling of combined municipal and industrial wastes.

Section 2

CONCEPT AND THEORY

Osmosis can be defined as the spontaneous passage of a liquid from a dilute to a more concentrated solution across an ideal semipermeable membrane which allows the passage of the solvent (water) but not the dissolved solids (solutes) (see Fig. 1). The transfer of the water from one side of the membrane to the other continues until the head or pressure (h) is large enough to prevent any net transfer of the solvent (water) to the more concentrated solution. At equilibrium, the quantity of water passing in either direction is equal, and the pressure (h) is then defined as the osmotic pressure of the solution having that particular concentration of dissolved solids.

If a piston is placed on the more-concentrated solution side of a semipermeable membrane (see Fig. 2) and a pressure, P, is applied to the solution, the following conditions can be realized: (1) P is less than the osmotic pressure of the solution and the solvent still flows spontaneously toward the more concentrated solution; (2) P equals the osmotic pressure of the solution and solvent flows at the same rate in both directions, i.e., no net change in water levels; (3) P is greater than the osmotic pressure of the solution and solvent flows from the more concentrated solution to the "pure" solvent side of the membrane. Condition (3), shown in Fig. 2, represents the phenomenon of reverse osmosis.

The osmotic pressure of a solution increases with the concentration of a solution. A rule of thumb, which is based on sodium chloride, is that the osmotic pressure increases by approximately 0.01 psi for each milligram/liter. This approximation works well for most natural waters. However, high-molecular-weight organics produce a much lower osmotic pressure. For example, sucrose gives approximately 0.001 psi for each milligram/liter.

OSMOSIS — NORMAL FLOW FROM LOW TO HIGH CONCENTRATION

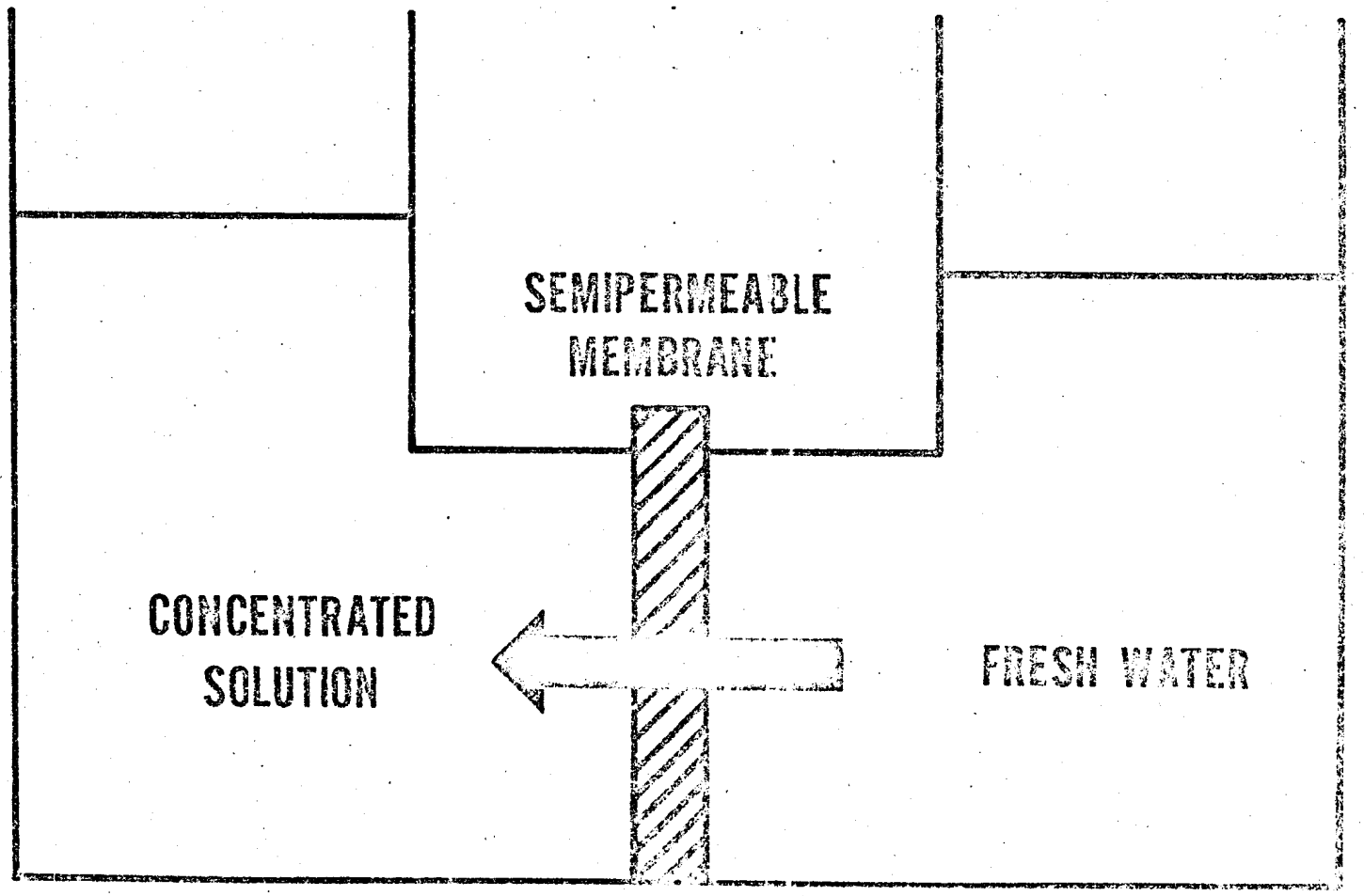


Fig. 1

REVERSE OSMOSIS – FLOW REVERSED BY APPLICATION OF PRESSURE TO HIGH CONCENTRATION SOLUTION

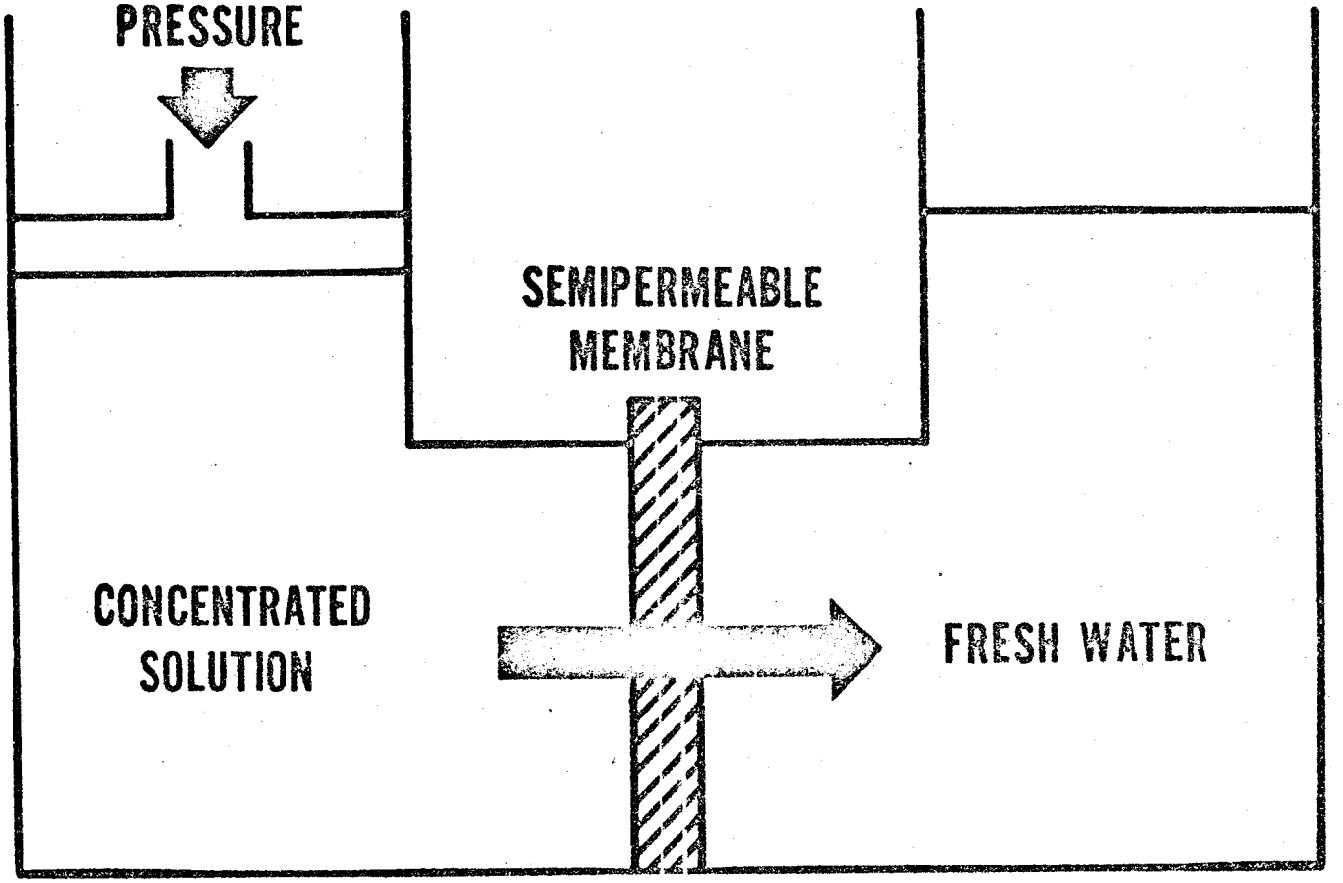


Fig. 2

A more accurate method is to use the van't Hoff equation. This is essentially the same as the ideal gas law:

$$\pi = \Phi \frac{n}{v} RT,$$

where π = osmotic pressure (atm),

Φ = osmotic pressure coefficient,

n/v = total concentration of ions or unionized species (moles/kg water),

R = gas constant = 0.082 liter-atm/degree-mol,

T = temperature (^{o}K).

Values for the osmotic pressure coefficient for many electrolytes are tabulated in the book, Electrolyte Solutions, by R. A. Robinson and R. H. Stokes (Butterworths, London). Although the osmotic pressure coefficient varies somewhat with concentration, for most electrolytes it is slightly less than 1, and it usually decreases with increasing concentration. If a value of 1 is used, the calculated osmotic pressure will normally be 5% to 15% higher than the observed osmotic pressure. (Table 1 shows the results of such calculations for some typical materials.) This allows a slightly conservative prediction of a unit's performance.

Several methods are available for measuring the osmotic pressure. It can be calculated from the depression of the vapor pressure of the solution and by depression of the freezing point. Several devices are commercially available for direct measurement of the osmotic pressure. These measure the pressure necessary to stop the flow of water through a membrane.

The procedure commonly used in evaluating the osmotic pressure of a solution is to measure the water flux through a module under operating conditions at several pressures. If a plot of water flux versus pressure is extrapolated to zero water flux, the intercept is the osmotic pressure, including any concentration polarization. During this procedure, care must

Table 1

Typical Osmotic Pressures at 25°C (77°F)

Compound	Concentration mg/liter	Concentration moles/liter	Osmotic Pressure psi
NaCl	35,000	.6	398
NaCl	1,000	.0171	11.4
$NaHCO_3$	1,000	.0119	12.8
Na_2SO_4	1,000	.00705	6
$MgSO_4$	1,000	.00831	3.6
$MgCl_2$	1,000	.0105	9.7
$CaCl_2$	1,000	.009	8.3
Sucrose	1,000	.00292	1.05
Dextrose	1,000	.00555	2.0

Note: Based on the above data for commonly present ionic species, a useful rule of thumb for estimating the osmotic pressure of a natural water supply requiring demineralization is 10 psi per 1,000 mg/l (ppm).

be taken to either maintain constant recovery or correct for the variation in concentration.

Attempting to measure the osmotic pressure of a solution directly by operating at a pressure just sufficient to obtain zero flow is impractical because the membranes are not perfect semipermeable membranes. This technique would measure the difference in osmotic pressure between the feed and the product water. At low pressures the salt rejection is relatively poor, so that a false osmotic pressure somewhat lower than the real value would be determined.

The basic behavior of semipermeable cellulose acetate reverse osmosis membranes can be described by two basic equations. The product water flow through a semipermeable membrane may be expressed as:

$$F_w = A(\Delta p - \Delta\pi), \tag{1}$$

where F_w = water flux (g/cm^2 sec),

A = water permeability constant (g/cm^2 sec atm),

Δp = pressure differential applied across the membrane (atm),

$\Delta\pi$ = osmotic pressure differential across the membrane (atm).

The salt flux through the membrane may be expressed as:

$$F_s = B(C_1 - C_2), \tag{2}$$

where F_s = salt flux (g/cm^2 sec),

B = salt permeability constant (cm/sec),

$C_1 - C_2$ = concentration gradient across the membrane (g/cm^3).

The water permeability and salt permeability constants are characteristic of the particular membrane which is being used and the processing which it has received.

An examination of Equations (1) and (2) shows that the water flux is dependent upon the applied pressure, while the salt flux is not. As the pressure of the feedwater is increased, the flow of water through the membrane will increase while the flow of salt remains essentially constant. It follows that both the quantity and the quality of the purified product increase with increased driving pressure.

The water flux clearly increases as the available pressure differential increases. It is also evident that the water flux decreases as the salinity of the feed increases – because the osmotic pressure contribution increases with increasing salinity. Further, as more and more water passes through the membrane as the feed passes through the unit, the salinity of the feedwater becomes higher and higher (concentrate). The osmotic pressure contribution of the concentrate becomes higher, and this then results in a lower water flux with increasing percent water recovery. Finally, since the salinity of the feed-concentrate increases with increasing product-water removal and the membrane rejects essentially a fixed percentage of the salt, the quality of the product water decreases with increasing percentage water recovery.

These points are well illustrated by Figs. 3 and 4. Figure 3 shows the variation in water flux for a reverse osmosis system operating under fixed pressure conditions as a function of percent water recovery, with feed salinity as a parameter. It is seen that the water flux decreases with increasing feed salinity and with increasing recovery percentages.

Figure 4 is a similar plot, except that product-water quality (salinity) is plotted versus percent recovery. Again, we see that the quality of the product water decreases with increasing feed salinity and increasing recovery percentages. It is noteworthy that for recoveries below about 75%, the reduction in water flux and water quality is fairly low, particularly on feeds of low salinity.

As noted above, the specific water and salt flows depend upon the water and salt permeation constants (A and B) of the membrane in question.

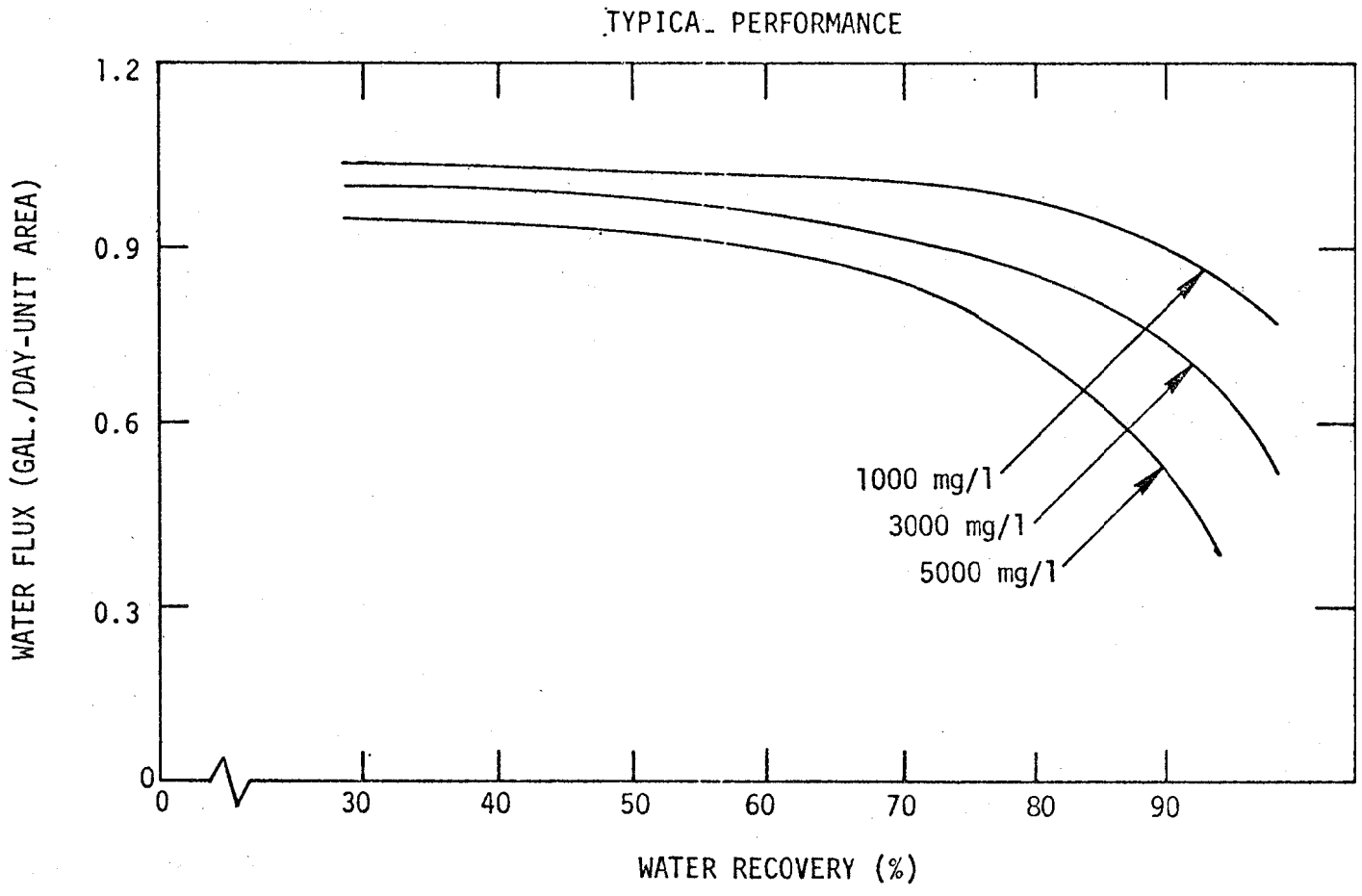

TYPICA_ PERFORMANCE
WATER FLUX (GAL./DAY-UNIT AREA)
1.2
0.9
0.6
0.3
0
0
30
40
50
60
70
80
90
WATER RECOVERY (%)
1000 mg/l
3000 mg/l
5000 mg/l

Fig. 3

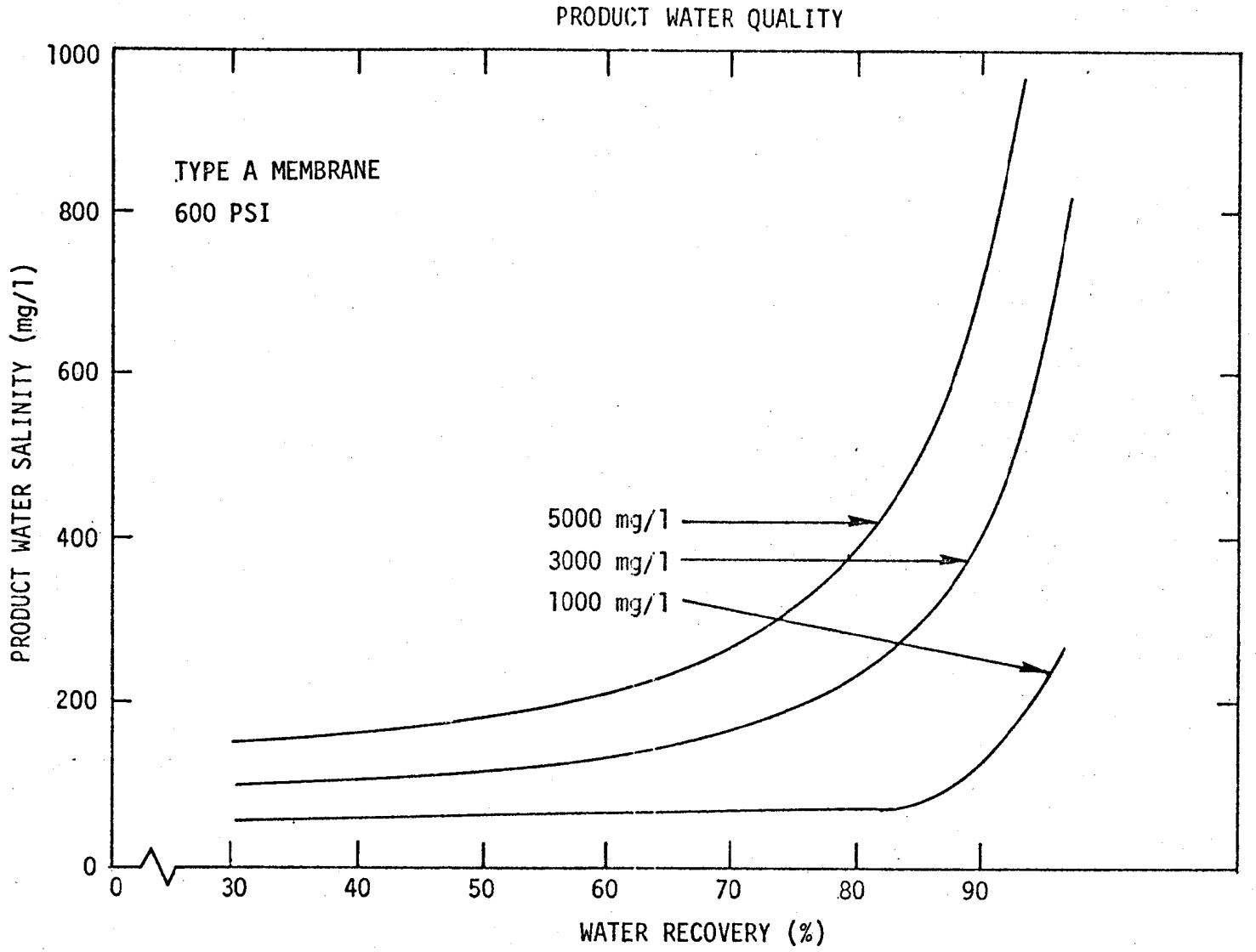

PRODUCT WATER QUALITY
TYPE A MEMBRANE
600 PSI
5000 mg/l
3000 mg/l
1000 mg/l
PRODUCT WATER SALINITY (mg/l)
1000
800
600
400
200
0
0
30
40
50
60
70
80
90
WATER RECOVERY (%)

Fig. 4

It is possible to prepare membranes over a range of specific water and salt permeation constants. When $A = 1.5 \times 10^{-5}$ g/cm^2 sec atm, the water flux rate corresponds to 13 gal/ft^2 day at 600 psi. As noted earlier, B is given in units of cm/sec.

An alternative method of expressing B, in more readily measurable terms, is:

$$B = \frac{C_p \times F_p}{a(\bar{c} - C_p)},$$

where C_p = salt concentration of product (g/cm^3),

F_p = flow of product (cm^3/sec),

$\bar{c}$ = average concentration of feed and concentrate (g/sec),

a = area (cm^2).

The percent salt rejection is defined by the equation:

$$\% \text{ S.R.} = 1 - \frac{C_p}{\bar{C}} \times 100,$$

where C_p = concentration (or conductivity) of the product,

$\bar{C}$ = average concentration (or conductivity thereof) of the feed and brine.

It then follows that a membrane exhibiting a water flux ratio of 13 gal/ft^2 day at 600 psi and 95% salt rejection will have a "B" value of 3×10^{-5} cm/sec.

References on Concept and Theory

1. Principles of Desalination by K. S. Spiegler, editor. New York: Academic Press, 1966, pp. 345 and following.

2. Desalination by Reverse Osmosis by V. Merten, editor. Cambridge, Massachusetts: MIT Press, 1966.

3. "Desalination by Reverse Osmosis," Process Synthesis and Optimization Case Problem by Roger W. Thompson and C. Judson King. Berkeley, California: University of California, October 1969.

4. The A-B-Seas of Desalting, U.S. Department of the Interior, Office of Saline Water, 1968.

5. Desalting – State-of-the-Art, Bulletin No. 134-69. State of California: Department of Water Resources, June 1969.

6. "Water Desalting Techniques – A Summary," Water and Wastes Engineering/Industrial, November 1969.

7. Reverse Osmosis by S. Sourirajan. London: Logos Press, 1969.

Section 3

APPLICABILITY OF REVERSE OSMOSIS TO PROPOSED SYSTEMS

This section discusses the applicability of reverse osmosis units for the various overall wastewater treatment systems being considered. The two types of systems proposed are: (1) the use of reverse osmosis units to handle the entire waste stream after primary or primary plus secondary treatment; and (2) the use of reverse osmosis units to handle only the reject stream from an ion exchange unit.

In system one, where the entire waste stream is being handled, the primary or primary plus secondary treatment is followed by filtration before the waste stream enters the reverse osmosis units. Under these conditions, and for a total dissolved solids (TDS) concentration of about 500 PPM, reverse osmosis is well suited to handle the wastewater in question. It is likely that electrodialysis will also be competitive with reverse osmosis for this waste stream and the optimum unit for such use may ultimately depend on the actual water quality parameters such as feedwater temperature, iron and manganese concentrations, and the relative importance of monovalent versus divalent ions in the feedwater.

In system two, where only the reject stream from an ion exchange unit is being handled, two different ion exchange systems are of interest. In the first system (the Chem-seps system), it is assumed that a product water with a TDS concentration of 10 PPM is produced and a waste stream of 3% of the total flow results. The reject stream will, therefore, have a TDS concentration on the order of 16,000 PPM. In addition, the overall reject stream could have high concentrations of calcium sulfate unless the two regeneration effluents were kept separate. Currently existing operational reverse osmosis unit are designed to handle up to 10,000 PPM TDS and can have problems with scale formation where calcium sulfate is present in high concentrations. Reverse osmosis is, therefore, not applicable for use at the present time in dealing with this type of

waste stream. Extensive research is, however, continuing in the search for reverse osmosis systems which are adequate to treat waters with TDS concentrations in excess of 10,000 PPM. It is expected that suitable systems will be available in the future.

The second ion exchange system (the Sanks system) is assumed to operate at 85% efficiency and yield a waste stream of about 10%. This reject stream will, therefore, have a TDS concentration of about 4000-5000 PPM. This effluent is within the limits of applicability of existing reverse osmosis (pretreatment may be required for pH control and wherever iron, manganese, and calcium sulfate concentrations are excessive). For these conditions, it is unlikely that electrodialysis will be a competitive process.

Section 4
DESIGN CRITERIA

Feedwater Parameters

Feedwater to a reverse osmosis system must meet certain criteria to permit satisfactory operation. The parameters of concern include TDS, calcium, sulfate, carbonate, iron, and manganese concentrations, pH, flow rate, temperature, suspended solids, turbidity, and oil or grease.

Current reverse osmosis units are designed to operate on a feedwater with a TDS concentration of from 500-10,000 PPM.* Units capable of operation above this range (e.g., for seawater, TDS of 35,000 PPM) are being developed and should be available in the future.

The reverse osmosis process divides the feedwater into two effluents — one, a purer product water and, the other, a more concentrated brine, the latter of which creates the possibility for scaling in the brine side where the solubility limits of calcium sulfate and/or carbonate are exceeded. This scaling problem may be encountered due to feedwater concentrations of calcium, sulfate, and carbonate, feedwater temperature, overall ionic strength of the feedwater, and the product recovery rate of the unit (which establishes the brine concentration). A conservative limit on the concentration of calcium in the brine is 1000 PPM (assuming, of course, that sulfate is present in at least a stoichiometric amount).

Under certain conditions, iron and manganese in the feedwater can oxidize and precipitate. This results in a coating on the reverse osmosis membranes, reducing the efficiency of the system. This coating occurs when the combined concentrations of iron and manganese exceed 0.3 PPM, although concentrations greater than this may be acceptable depending on

*For operation at the low end of this range, electrodialysis may be a competitive process, depending on feedwater temperature and the relative concentrations of monovalent and divalent ions.

the valence of the ions present and the pretreatment of the feedwater.

The cellulose acetate membranes used in reverse osmosis are subject to hydrolysis and the optimum range of operation is for pH between 4 and 5.5. The rate of hydrolysis is a critical factor in membrane life. Hydrolysis rate is a minimum where feedwater pH is about 5, but with pH values below as well as above 5, the rate of hydrolysis is increased.

In general, the water flux through reverse osmosis membranes increases with increasing feedwater temperature. Another way of expressing this is that the required membrane area for a given flow rate decreases for temperatures above a standard of 70°F. The required area decreases or increases by 1.8% for every degree above or below 70°F. Operation with feedwater temperatures up to 85°F is acceptable. Temperatures of up to 100°F can be tolerated for short intervals. Operation above 100°F for any length of time or extended operation above 85°F may lead to decreased membrane life.

Feedwaters with high turbidity can foul membrane surfaces and, for optimum operation, turbidity should be kept below one Jackson Turbidity Unit and all particulate matter larger than about 25-100 microns should be eliminated.

Oil or grease in the feedwater can also foul membranes by forming a coating. Oil should, therefore, be removed wherever present to avoid this possibility.

Performance Parameters

The salt rejection factor for the reverse osmosis membrane is assumed to be 95% and the flux is 15 gfd (gallons per square foot per day). The 95% rejection factor is the efficiency with which the membrane operates at any point within the desalting system. Where the feedwater first enters the system, the product water at that point has a concentration of total dissolved solids equal to 5% of that for the feedwater (95% rejection). At the very end of the system, however, the water contacting the membrane has

become concentrated by the desalting which has taken place up to that point. This water is now about equal in quality to the brine discharge and the product water produced at this point will have a much higher concentration of total dissolved solids than the product water produced at the beginning of the system. The 95% rejection factor still applies, but the "feedwater" at this point is greatly different from the feedwater entering the system. To determine the overall product water quality, the 95% rejection factor is applied to a figure corresponding to the arithmetic average of the beginning feedwater and brine concentrations.

The membrane flux of 15 gfd is applicable only for operation at the nominal or design system pressure of 600 psi. If operation at lower system pressures is required or desirable, the membrane flux is calculated using the following relationship:

$$J = K\ (P_s - P_o),$$

where J = membrane flux, gfd,

K = a constant specified as 0.4 gfd/atm of pressure,

P_s = system pressure, atm,

P_o = osmotic pressure, atm.

The osmotic pressure is that for the feedwater which can be assumed to be 0.00082 times the concentration of total dissolved solids in mg/l.

Water recovery rates for reverse osmosis units should be as high as possible to maximize product water and minimize brine. At recovery rates above about 80%, the water flux declines and product water salinity increases. For these reasons, 80% product water recovery rate is about the maximum for practical usage.

Section 5

PRETREATMENT REQUIRED

Just as for other more conventional processes, the performance of reverse osmosis can be adversely affected by certain characteristics of the feedwater. This section will discuss the pretreatment methods required to avoid such losses of efficiency. The basic problems to overcome are membrane hydrolysis, fouling of membrane surfaces, and scale formation.

Membrane hydrolysis is minimized by proper pH control for the feedwater. The optimum desired pH is from 4 to 5.5 and this is achieved by acid or base addition as necessary.

Membrane fouling can occur by suspended solids, oil, and iron or manganese precipitates. Suspended solids can be removed from the feedwater by filtration. To accomplish this filtration, pressure and gravity sand filters and diatomaceous earth filters are suitable, but where the particulates approach or are colloidal, chemical treatment (coagulation-flocculation) and filtration may be required. Oil can be removed by absorption techniques. Coagulation with chemical flocculants and removal by sedimentation and filtration is a suitable method. Where iron and manganese are present, the problems of oxidation and precipitation can be minimized by avoiding exposure of the feedwater to air and by pH control (similar to that used to protect membranes from excessive hydrolysis). In extreme cases, pretreatment with potassium permanganate for iron and manganese removal may be required.

Scale formation can occur by precipitation of calcium sulfate or carbonate if the solubility limits of these compounds are exceeded. Pretreatment includes pH control and the addition of sodium hexametaphosphate.

Section 6

EFFLUENT QUALITIES

The application of reverse osmosis to the proposed design schemes is considered in two modes: (a) reverse osmosis as a method of handling the rejection from ion exchange systems and (b) reverse osmosis as a separate tertiary treatment step. For both cases, input information is based on the following design data.

Area:	Boston	Lowell-Lawrence-Haverhill	Winnipesaukee
Influent Characteristics:			
Design Flow	500	121	10
BOD	175	215	164
Suspended Solids	260	245	158
COD	260	320	275
Nitrogen	25	15	30
Phosphorous	12	10	15
Total Dissolved Solids	300	335	300

Specific effluent criteria for the reverse osmosis units will be developed in subsequent portions of the report as part of the overall evaluation of the reverse osmosis unit as it is employed in a given design scheme.

Section 7

OPERATIONAL DATA OF REVERSE OSMOSIS PLANTS

Commercial desalting plants using reverse osmosis which were in operation or under construction as of January 1, 1969, are shown below.

Location	Capacity (1000 gals/day)	Feedwater	Use
San Diego, California (River Valley)	50	Inland saline or brackish water	Irrigation
Plains, Texas	100	Inland saline or brackish water	Municipal water supply
Yotuata, Israel	53	Inland saline or brackish water	Municipal and industrial water supply

Field tests using reverse osmosis test units have been and are continuing to be conducted by the Office of Saline Water, U.S. Department of the Interior. Details of these field tests are given in the following table.

Table 2

Location	Type of Unit	Capacity (1000 gals/day)	Feedwater	Use
Ft. Morgan, Colo.	Tubular Plate and frame Spiral-wound Hollow fine-fiber	1 10 2 6	Brackish well water	Municipal water supply
Grand Junction, Colo.	Tubular Spiral-wound	3 2.5	Irrigation-return water, brackish well and surface water	Irrigation and salinity control
Las Vegas, Nev.	Tubular Spiral-wound	-- 2.5	Surface water polluted with industrial waste and municipal sewage	Not specified
Roswell, N.M.	Plate and frame Spiral-wound Tubular	10 10 1 & 3	Brackish well water	Municipal water supply
Webster, S.D.	Hollow fine-fiber	6	Brackish well water	Municipal water supply
Denver, Colo.	Tubular	3	Brackish well water	Municipal water supply
Norton, W.V.	Spiral-wound	10	Acid mine water	Pollution control

Based on the tests reported in Table 2, it is apparent that various types of reverse osmosis units are capable of successfully treating a wide range of feedwaters. The solutions to a variety of operational problems which were encountered during these field tests contributed to the advancing state-of-the-art of reverse osmosis desalting.

Section 8

CAPITAL AND OPERATING COSTS

In order to calculate costs for the various systems, a set of technical and economic criteria must be delineated. These are indicated in the following paragraphs.

Technical Criteria

Based on the current state-of-the-art, the following performance characteristics for reverse osmosis systems have been used for making calculations:

1. Maximum product recovery rate of 75%.
2. Salt rejection of 95% at operating pressure.
3. Membrane performance of 15 gfd (gallons per square foot per day).
4. Operating pressure of 600 psi.

Economic Criteria

1. Interest rate for financing, 6%.
2. Years for amortizing equipment, 30.
3. Land costs for locating plant, $2,000/acre.
4. Labor costs are equivalent to U.S. average.
5. Basic costs are derived from the Office of Saline Desalting Cost Calculating Procedures (OSW R&D Report 555, February 1970) which have been updated using the following:

   A. Building cost correction factor of 1.11 (applicable for first quarter, 1971).

   B. Labor cost correction factor of 1.08 (applicable for fourth quarter, 1970).

C. Additional capacity scaling information has been taken from the 1969-1970 Saline Water Conversion Report.

Cost Calculations

Desalting costs for reverse osmosis systems are generally presented in terms of cents per thousand gallons of product water. This method of expressing costs is important to remember in examining such costs since the intake and product water quantities for reverse osmosis are significantly different.

Costs for reverse osmosis treatment as part of the various overall treatment systems in the three areas of interest are given below.

	(Costs in Cents/1000 gal)		
	Boston	Lowell-Lawrence-Haverhill	Winnipesaukee
System One (handling the rejection from ion exchange systems)			
A. Chem-seps system*	48.8	57.4	96.3
B. Sanks system	32.8	39.9	54.6
System Two (as a separate tertiary treatment)	31.5	31.5	40.6

*This will be an advanced reverse osmosis system and it has been assumed that costs for such a system will be 25% greater than those of a conventional system.

Corresponding capital and operating costs for these systems are given below.

	Boston	Lowell-Lawrence-Haverhill	Winnipesaukee
System One			
A. Capital Cost	$ 8,100,000	$ 1,880,000	$ 344,000
Annual Operating Cost	555,000	284,000	24,000
B. Capital Cost	12,400,000	4,110,000	533,000
Annual Operating Cost	3,490,000	764,000	63,000
System Two			
Capital Cost	116,000,000	28,000,000	3,500,000
Annual Operating Cost	33,700,000	8,190,000	630,000

Owing to the lack of information on reverse osmosis system costs for capacities above 50 MGD, it has been assumed that costs vary linearly with capacity above 50 MGD.

Section 9

SAMPLE CALCULATIONS

It has been assumed that a 75% product water recovery rate will apply for the reverse osmosis unit of interest. In this case, there will be three parts of product water for every four parts of intake water, and one part of brine will result.

For the 75% product recovery rate and 95% salt rejection rate, the TDS of the product will be 0.1163 times that of the intake TDS and the TDS of the brine will be 3.651 times that of the intake TDS.

Capacities for product water and brine streams for the various systems in the three areas of concern are as follows.

	Capacity in MGD (Product/Brine)		
	Boston	Lowell-Lawrence-Haverhill	Winnipesaukee
System One			
A. Chem-seps system	11.25/3.75	2.7/0.9	0.225/0.075
B. Sanks system	37.5/12.5	9.1/3.0	0.75/0.25
System Two	375/125	91/30.3	7.5/2.5

Product and brine qualities for these systems are given below (qualities for all cities are about equal).

	Effluent TDS	
	Product	Brine
System One		
A. Chem-seps system	1,900	58,000
B. Sanks system	500	16,000
System Two	58	1,800

An example of the breakdown of capital and operating costs is given on the following page.

System Two (A)
Capacity (11.25 MGD)
Boston

REVERSE OSMOSIS PROCESS
SUMMARY WORKSHEET

Cost Elements	$\$10^3$	Annual Cost $	Water Costs ¢/kgal
Capital Costs			
1. Plant and Equipment	6,100	443,000	12.0
2. Feedwater Pretreatment			--
3. Feedwater Supply			--
4. Water Transmission			--
5. Brine Disposal			--
Total Capital Costs	6,100	443,000	12.6
Operation and Maintenance Costs		Annual Cost $	
6. Operating and Maintenance Labor			
a. Plant and Equipment		158,000	4.5
b. Feedwater Pretreatment			--
c. Feedwater Supply			--
d. Water Transmission			--
e. Brine Disposal			--
Total Operating and Maintenance Labor			4.5
7. Other Operation and Maintenance Costs		Annual Cost $	
a. Payroll Extras (15% of 6a)		23,700	0.6
b. General and Administrative Overhead (30% of 6a + 7a)		54,500	1.6
c. Supplies and Maintenance Materials		61,000	1.7
d. Membrane Assembly or Replacement Tubing			13.2
e. Chemicals			4.0
f. Fuel or Steam			n.r.
g. Electric Power			
Plant and Equipment (for $.01/KWH electric cost)			10.6
Feedwater Supply			--
Water Transmission			--
Total Other Operation and Maintenance Costs			31.7
Total Operation and Maintenance Costs			36.2
Total Water Cost (Total Capital Plus O. & M. Costs)			48.8

n.r. - none required

Section 10

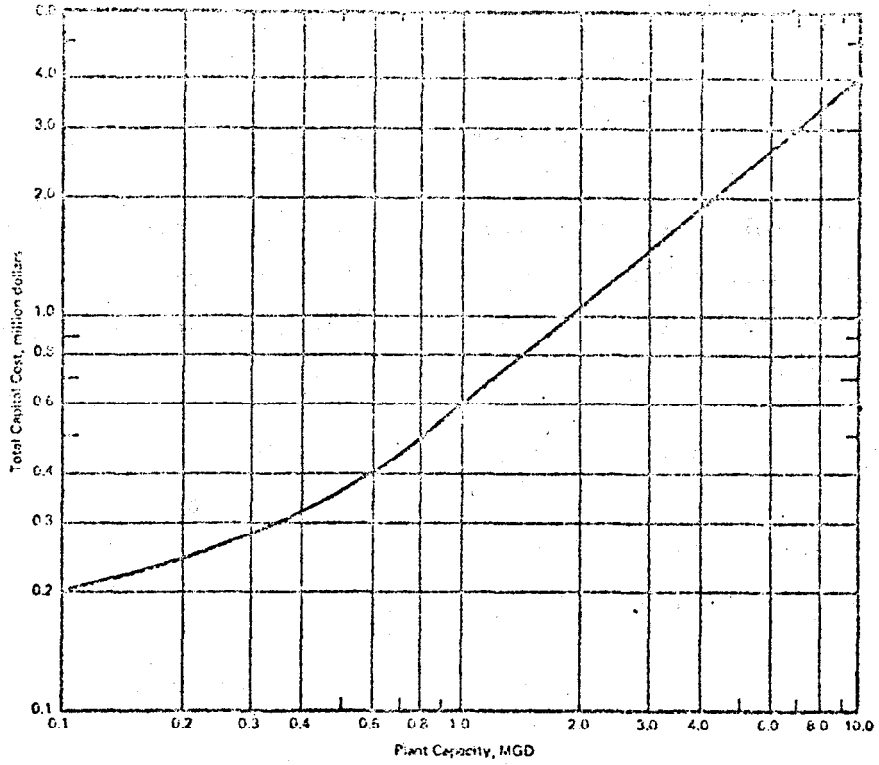
SCALE-UP FACTORS

The basic scaling curves for capital and operating labor costs applying to reverse osmosis systems are given in Figs. 1 and 2. This information is current for December 1969 prices and must be updated as required with cost correction factors.

The information from Figs. 1 and 2 must be blended with addition costing data from the 1969-1970 Saline Water Conversion Report when capacities above 10 MGD are required and the appropriate data is given below.

Capacity (MGD)	Overall Water Costs (¢/1000 gal)
1	57
10	42
50	34

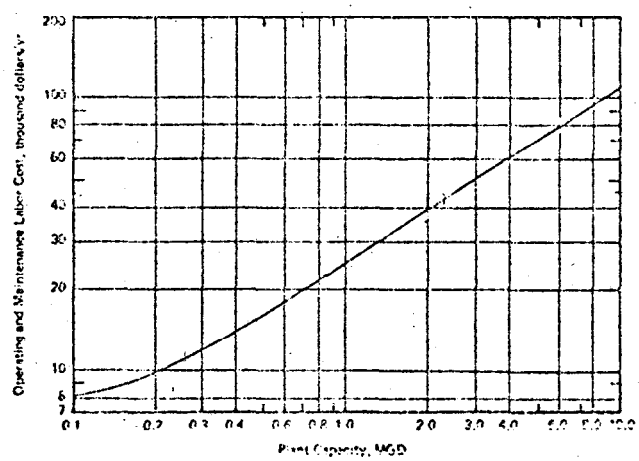
Information for systems with capacities above 50 MGD are not presently available, so it has been assumed that costs are linearly related with capacity in this range.



Source: Office of Saline Water and Southwest Research Institute

Figure 1

REVERSE OSMOSIS PLANT CAPITAL COSTS



Source: Southwest Research Institute

Figure 2

REVERSE OSMOSIS OPERATING AND MAINTENANCE LABOR COSTS

Section 11

PROCESS DEPENDABILITY

Many field tests of reverse osmosis units have been conducted in which the units performed satisfactorily for over one year of operation. Current costing information is based on a membrane life of one year and this appears to be justified at this time.

As field tests and research continue, it is expected that a membrane life of about three years will be achieved in the future. For such projected technology, the costs of reverse osmosis treatment will be greatly reduced.

Section 12

SUMMARY

The following table gives a summary of selected parameters and expected ratings for future target dates.

	Target Year		
	1975	1990	2020
Scale-up	10-50 MGD	50-500 MGD	500 MGD +
Dependability over Time	1-3 years	3-5 years	5+ years
Flexibility (in relation to other processes)	Limited	Better	Good
Tolerance to Influent Variability	Good	Better	Better
Effluent Variability	Some with age	Better	Better
Ancillary Processes Required	Some	Few	None
Costs (¢/1000 gal)	30-50	25-40	15-30
Process Efficiency	Good	Better	Better
Operational Difficulty	Some	Better	Better

Appendix

BIBLIOGRAPHY OF REVERSE OSMOSIS APPLIED TO WASTEWATER TREATMENT

Applied Science and Technology

1. "Multi-purpose Desalination by Reverse Osmosis," Engineering, 202:550, September 23, 1966.

2. "Reverse Osmosis for Water Reclamation," WPCF J, 38:397-8, March 1966.

3. "Reverse Osmosis Now Gets Tryout for Water Renovation," Chem Eng, 72: 102, November 22, 1965.

4. "Reclaimed Wastewater for Santee Recreational Lakes," WPCF J, 38: 1310-18, August 1966.

5. "Reverse Osmosis as a Concentration Technique," Anal Chem, 38:351-3, February 1966.

6. "Aerojet Moving into Water Purification," Tech W, 20:26-8, March 6, 1967.

7. "Desalination by Reverse Osmosis," Ind & Eng Chem Fundamentals, 6:2-12, February 1967.

8. "Reverse Osmosis Techniques Star at Water Conference," Tech W, 20:21-2, May 29, 1967.

9. "Cellulose Acetate Membranes for Waste Water Purification," Mod Plastics, 45:141-4+, May 1968.

10. "Engineering for Pure Water; Reverse Osmosis," Mech Eng, 90:104-10, April 1968.

11. "New Product Concept for Improvement of Industrial Waters," WPCF J, 40:439-46, March 1968.

12. "Semipermeable Hollow Fibers Remove Pollutants from Water in Du Pont Tests," Chem & Eng N, 45:66, October 23, 1967.

13. "Water Purifier Operates on Reverse Osmosis Process," Mech Eng, 90:66, September 1968.

14. "Improved Water Quality," Chem Eng Prog, 65:62-8, January 1969.

15. "New Technique for Purifying Water, Reverse Osmosis Process," Plant Eng, 22:92-3, December 12, 1968.

Applied Science and Technology (continued)

16. "Reverse Osmosis Helps Combat Water Pollution," Product Eng, 40:18, January 27, 1969.

17. "Reverse Osmosis Purifier Water," Food Eng, 40:121, December 1968.

18. "Water Pollution Control, Advanced Treatment Processes," Chem Eng, 75:95-9, October 14, 1968.

19. "Water Pollution, Removing Solubles from Waste Water," Can Chem Process, 53:49-51, October 1969.

20. "Water Renovation by Reverse Osmosis," Chem & Ind, 297-301, February 28, 1970.

21. "Waste Water Treatment with a Profit Potential," Pub Works, 101:82-4, April 1970.

Pollution Abstracts

1. "Reverse Osmosis for Wastewater Treatment," Industrial Water Engineering, 7(10):16-19, October 1970.

2. "Reverse Osmosis Process for Water Reuse Application," 1969 Symposium Sponsored by American Institute of Chemical Engineers, New York City, pp. 140-7. In AICE Chemical Engineering Progress Symposium Series.

3. "Reverse Osmosis Field Testing on Acid Mine Waters at Norton, West Virginia," U.S. Office of Saline Water, Washington, D.C., Research and Development Progress Report No. 586, 49 pp., July 1970.

4. "Development of an Improved Tubular Reverse Osmosis Module for Water Treatment," U.S. Office of Saline Water, Washington, D.C., Report No. 576, 121 pp., December 1970.

5. "One-way Filth Strainer," New Scientist, London, 46(696):58-60, April 9, 1970.

6. "Rivers as Resources," New Scientist, London, 46(696):51, April 9, 1970.

7. "Ultrafiltration, Pollution Engineering, 2(1):30-33, March-April, 1970.

8. "Reverse Osmosis Can Cut Cost of Water Treatment," Industrial Water Engineering, 7(3):25-30, March 1970.

Selected Water Resources Abstracts

1. "Water Reclamation by Reverse Osmosis," W70-02058 05D W3 06.

2. "Development of Water Purification Systems Based on Reverse Osmosis," W70-04086 03A W3 11.

3. "Reverse Osmosis Process for Water Reuse Application," W70-07389 05D W3 18.

4. "Membrane Processes Gain Favor for Water Reuse," W70-08448 05D W3 20.

5. "Reverse Osmosis Can Cut Cost of Water Treatment," W70-08851 05F W3 21.

TERTIARY TREATMENT BY FILTRATION IN THE MERRIMACK RIVER BASIN

by

E. Robert Baumann, P.E.
Professor of Civil Engineering
Iowa State University

To

Anderson-Nichols and Company, Inc.
150 Causeway Street
Boston, Massachusetts 02114

LIST OF TABLES

Table 1. Variation in monthly wastewater flow rates expressed as percentages of the average yearly flow.

Table 2. Variation in daily flow rate expressed as percentages of the average weekly flow rate.

Table 3. Variation in hourly flow rate expressed as percentages of the average daily flow.

Table 4. Typical filtration plant characteristics for study areas.

Table 5. Summary, Granular Filters for Wastewater Filtration (a literature summary).

Table 6. References cited in Table 5.

Table 7. Tertiary filtration plant locations.

Table 8. Capital costs of tertiary filtration plant.

LIST OF FIGURES

Figure 1. Operational optimum filter design.

Figure 2. Filter configurations.

Figure 3. Influent-control, constant rate filter.

Figure 4. Operating cycles of influent-control, constant-rate filter.

Figure 5. Declining-rate filter.

Figure 6. Ranges in typical 24-hr wastewater flow pattern.

Figure 7. Capital costs for tertiary treatment filters.

Figure 8. Least cost design characteristics of filters for iron removal (2).

Figure 9. Filter characteristics, phase A.

Figure 10. Phase A, Q = 6gpm/sq ft., 20-in. deep media.

Figure 11. Phase A, 0.92mm sand, Q = 6gpm/sq ft., variable media depth.

Figure 12. Filter characteristics, phase B.

Figure 13. Phase B, Q = variable, 20-in. deep media.

Figure 14. Phase B, 0.92mm sand media, Influent SS = 22 mg/l, Q = variable.

Figure 15. Filter characteristics, phase C.

Figure 16. Phase C, Q = 2gpm/sq ft., 20-in. deep media.

Figure 17. Phase C, 1.84mm coal - 0.75mm sand media, Q = 2gpm/sq ft., depth variable.

Figure 18. Phase C, Q = 4 gpm/sq ft., 20-in. deep media.

Figure 19. Phase C, 1.84mm coal - 0.75mm sand filter, depth variable, Q = 4gpm/sq ft.

Tertiary Treatment by Filtration
in the Merrimack River Basin

1. DEFINITION OF THE PROBLEM

This paper deals with the achievement and maintenance of water quality in a major river basin in the United States, namely the Merrimack River Basin of Massachusetts and New Hampshire. In order to better understand the specific problems of this basin, three different regions of widely differing character have been selected for study. These are the:

Winnipesauksee Study Area, 1990 Mean Waste Flow of 9.72 MGD

Lowell-Lawrence-Haverhill Study Area, 1990 Mean Waste Flow of 121 MGD

Boston Study Area, 1990 Mean Waste Flow of 500 MGD

In these study areas, the wastewater collection systems range from old combined sewers to recently constructed separate systems. The domestic waste components vary from small volumes in some areas of the basin to relatively large volumes in the major urban complexes. The objective of this report is to consider the various factors affecting the design of a feasible waste water treatment system which will provide "maximum" improvement of water quality. This report is addressed to a consideration of the applicability of FILTRATION processes in a tertiary waste treatment scheme.

In preliminary discussions, the objectives of the waste treatment were deemed to include:

- Reduction of BOD_5 at 20°C to about 2 mg/1.
- Reduction of Suspended Solids to level of 2 mg/1 (Turbidity Units of 4-5).*
- Reduction of COD to a level of 4-5 mg/1.

---

* USPHS drinking water standard sets allowable turbidity units at 5.

- Reduction of total Nitrogen and phosphorous to a level less than 1 mg/l.

- Reduction of total dissolved solids to a level equivalent to river water, or about 100 mg/l.

In order to accomplish this, two treatment schemes have been proposed:

- Scheme 1 Conventional biological treatment followed by coagulation, up-flow sedimentation, adsorption in expanded carbon beds, denitrification, filtration, ion exchange, and membrane concentration of the brine solution from ion exchange for ocean disposal.

- Scheme 2 Physical-chemical treatment involving coagulation, up-flow sedimentation, ammonia removal, adsorption in expanded carbon beds, filtration, ion exchange and membrane concentration of the brine solution from ion exchange for ocean disposal.

Both schemes would involve solids disposal of all solids concentrated in the treatment schemes. In both schemes, the partially treated wastewater applied to the filters would have undergone treatment on expanded carbon beds and would have approximately the following characteristics:

Filter Waste Characteristics

Waste Characteristic	ON FILTERS	OFF FILTERS
BOD_5, mg/l	5	2
SS, mg/l	18	2 (4-5TU)
COD, mg/l	10	5
N, mg/l	1	1
P, mg/l	1	1
TDS, mg/l	275	275

This report considers the factors affecting the types of filters -- and their design characteristics -- which might be used for providing the desired filtered water quality.

2. CONCEPT AND THEORY

a. Design considerations

If water containing suspended solids is passed through a layer of porous media, some of the suspended and colloidal material are partially removed. This process is called FILTRATION and its efficiency and cost is a function of:

- the concentration and characteristics of the solids in suspension (particle size distribution, surface characteristics, organic vs. inorganic, etc.),
- the characteristics of the filter media and other filtering aids used (particle size distribution, surface characteristics, etc.),
- the characteristics of the solids in solution in the water filtered, and
- the characteristics of the filter and its method of operation.

The criteria of importance which must be considered in design involve finding:

- the OPERATIONAL OPTIMUM filter design characteristics, and
- the ECONOMIC Optimum filter design.

During a filter run, the head loss across the filter media will increase due to the accumulation of solids within the filter media. When this head reaches the limit set by the hydraulic conditions of the design, the filter run must be stopped. Figure 1 indicates that as the head loss increases during a filter run, the filtrate quality also changes, tending to rise in value as time proceeds. Although the filter could be designed to produce a satisfactory filtrate quality during the early stages of the run, there will come a time when the filtrate quality will become unsatisfactory and the filter run will have to be terminated due to solids breakthrough above the critical concentration of suspended solids, (Cc), permissible. A filter OPERATIONAL

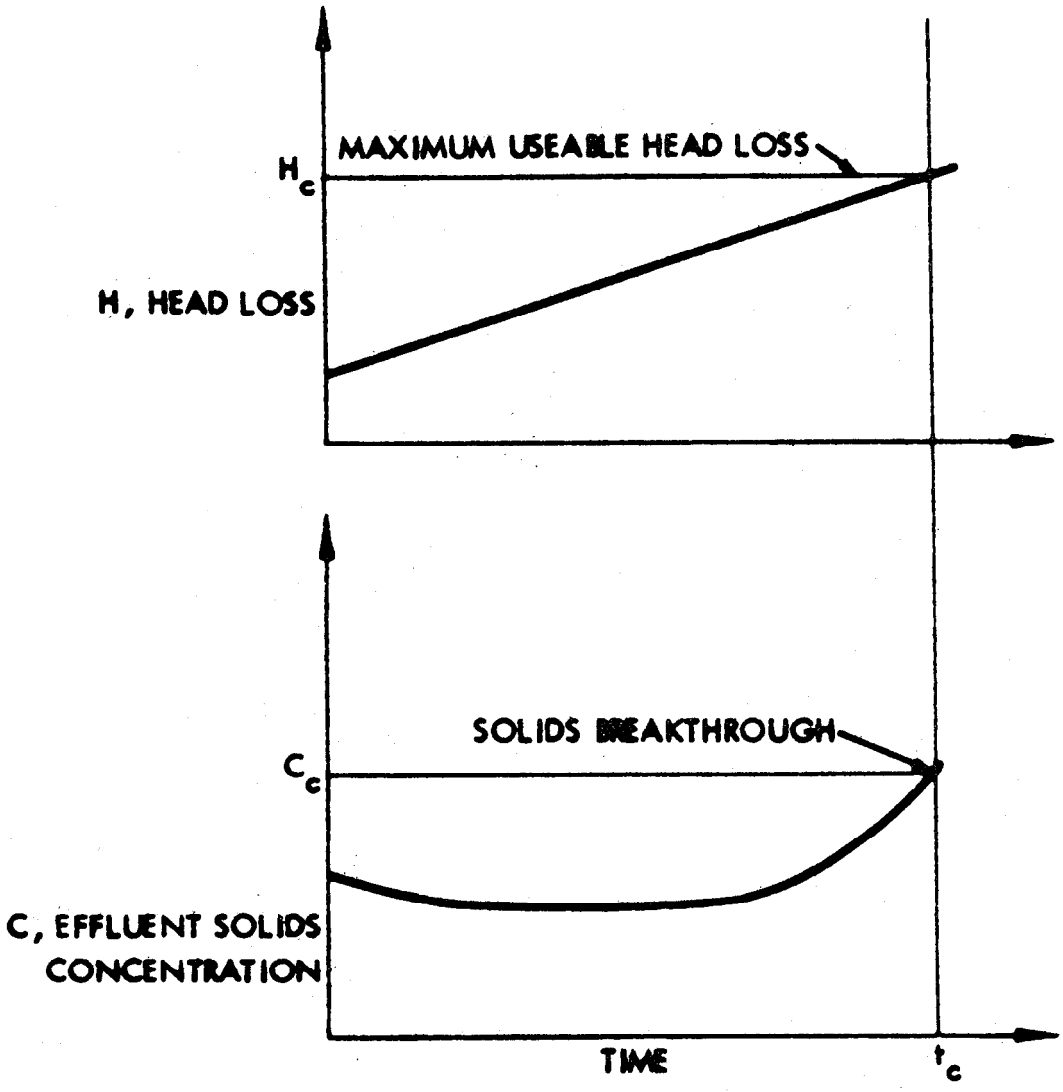


FIG 1. Operational optimum filter design.

OPTIMUM condition occurs when both the head loss and effluent quality reach their respective critical values (Hc and Cc) at the same time.

To achieve an OPERATIONAL OPTIMUM, many alternative designs are possible which can produce "equivalent performance". Two or more filters may be said to provide equivalent performance when they produce the same quality and quantity of filtered water from the same water source during the same time period. Of the equivalent performance filters, however, only one will produce water at the least total cost per million gallons. Current trends indicate that we are approaching the time when we will be able to design filters to provide both OPERATIONAL and LEAST COST OPTIMUMS.

Operational optimum design requires consideration only of the:

- Media size,
- Media depth,
- Flow rate, and
- Hydraulic conditions affecting the head loss.

In a least cost optimum design, the cost effects of the filter structure, hydraulic appourenances, energy, labor and maintenance requirements, and the size of the total filter plant also have to be considered. Ives (1) and Huang and Baumann (2) have both recently considered in detail the data needed for design of OPERATIONAL LEAST COST OPTIMUM designs for granular media filters suitable for tertiary waste treatment applications. To accomplish this, three elements are necessary:

- A method (mathematical model) to predict the performance of the filters to determine combinations of the filter design variables which will provide the same quality of effluent under operational optimum conditions.
- A filtration system whose first cost, operating cost, and maintenance cost can be predicted with reasonable accuracy.

- A computer program which can provide a list of operational optimum filter designs and predict the one design which will provide the treated water at least cost.

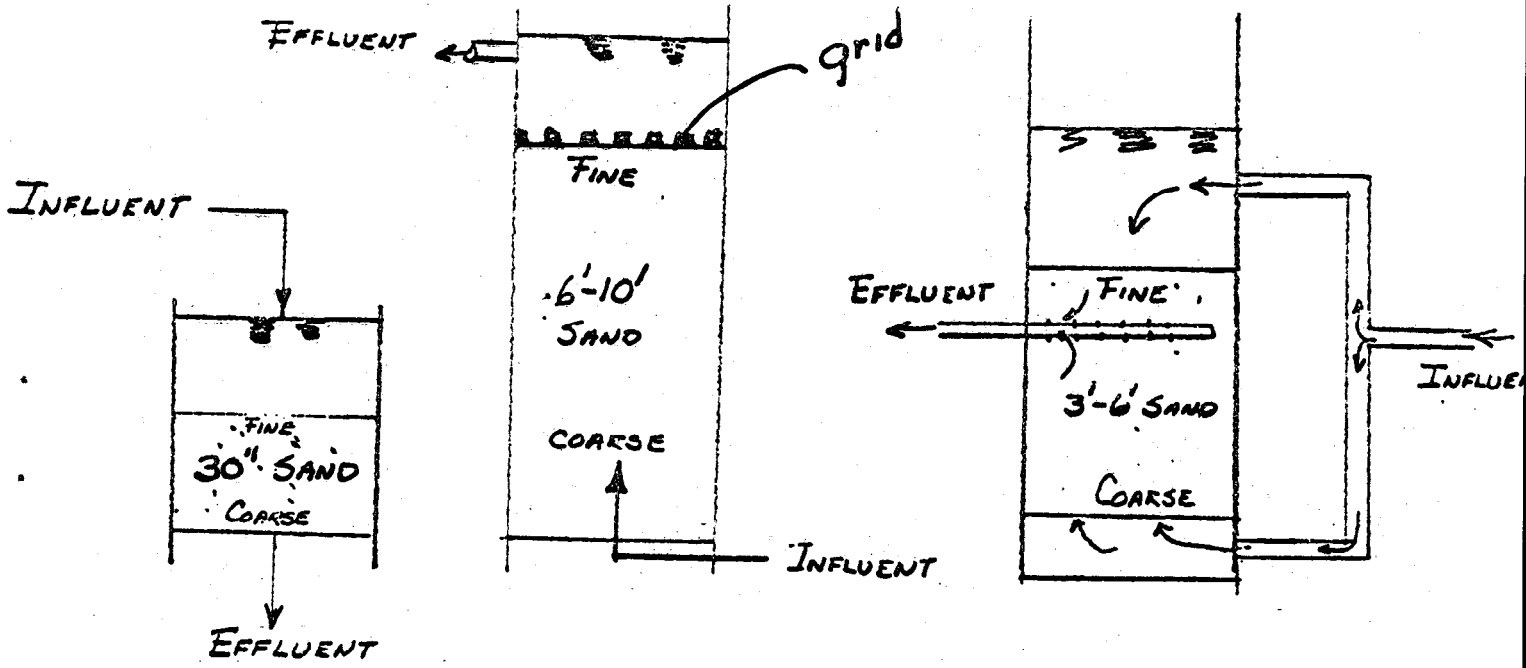
To date, no such mathematical model -- or even filter operating experience with wastewater from expanded carbon beds -- is available for consideration in the tertiary filtration application considered herein. Accordingly, it would appear that prior to detail design of filters for the three study basins, the required operational optimum least cost mathematical model be developed by:

- Running pilot plant studies using a water pretreated as proposed in the two flow schemes.
- Building the complete treatment plant for Winnipesaukee based on pilot plant design with enough filter flexibility for perfecting the model using full-scale plant filters and the actual waste to be treated.

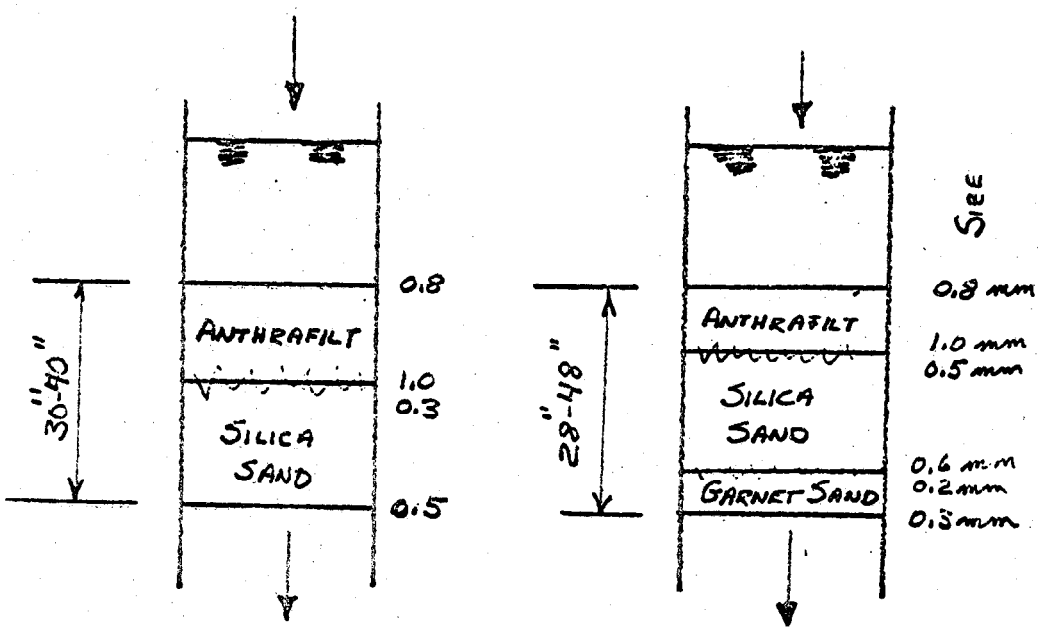
The analysis of the data collected could be accomplished using the techniques of Hsuing-Cleasby (3). Until such data are available, the design requirements of tertiary filtration plants for the two schemes of treatment recommended in this report must be based on the author's personal wastewater filtration experiences and on the experience of others.

b. Filter configurations

Figure 2 shows several filter configurations which have been used extensively in the past few years in water filtration. The granular filter originally was used for potable water filtration with a single media (effective size = 0.5 mm, 1.5-1.8 uniformity coefficient) and downward filtration through the finest media which collected at the top of the filter at low filtration rates (2 gpm/sq ft). In recent years, design trends are to use of higher filtration rates (3 to 5 gpm/sq ft), deeper media, and coarser



SINGLE-MEDIA FILTERS



DUAL MULTI

MEDIA FILTERS

FIG 2 FILTER CONFIGURATIONS

media. To better distribute solids within the filter media, several filter configurations have been tried:

- (Fig. 2b) Using upflow filtration through a relatively deep and coarse filter media.
- (Fig. 2c) Using a filtered water collection device within the filter media and bringing water in from both the bottom and top of the filter media.
- (Fig. 2d, e) Using dual or multi-media with downward flow as in conventional filters.

British practice in tertiary treatment tends to use the Fig. 2b configuration (4). This configuration has several advantages:

- Filtration can take place from coarse to fine media using a single media for better solids distribution. Therefore, a coarser effective size with a larger uniformity coefficient media can be used.
- Effective backwash time is less since drain-down time is significantly less.
- Raw water is used for backwashing reducing somewhat the amount of water which must be filtered twice.
- There is no necessity for having high walls on the filter tank in order to build up the necessary static head required for filtration.

The chief disadvantages of this configuration are:

- The need for a bar screen configuration within the sand near the surface to retain the media in place against the upward force exerted during filtration.
- There is a tendency for the media retaining straining nozzles used to deliver raw water to the filters to clog when suspended solids in the filter are high.

American practice tends to the use of two or three media as in Figs. 2d and e. Usually, a coarse uni-size anthrafilt (Sp. Gr. = 1.4-1.6) is used on top of a finer uni-size silica sand (Sp. Gr. = 2.65). Occasionally, a still finer uni-size garnet sand (Sp. Gr. = 3.5$\pm$) is used at the bottom. Dual or multi-media filters have the following advantages:

- Filter design follows current American practice and permits production of the desired quality of effluent with reasonable run lengths.

- No filter media restraining grids or underdrain clogging problems are encountered.

They also have the following disadvantages:

- Dual or multi-media filters require relatively deep filter cells to provide the required filtering head without creating negative head conditions in the filter.

- Filtering down time in preparation for backwashing is significant and extends the out-of-service time for backwashing to about 30 minutes minimum.

- Use of dual or multi-media filters creates a need for more care in backwash design to prevent loss of media or excessive intermixing of the media.

Although both Fig. 2b and Fig. 2d and e filter configurations have applications in tertiary wastewater filtration, only the latter filter configuration will be considered for use in this report.

c. Method of operation

Two methods of filtration are most applicable for tertiary wastewater filtration, constant rate with influent rate control or declining rate filtration. Rate control for constant-rate gravity granular filters has been achieved for many years by effluent control systems. The valves used as rate controllers frequently do not function properly and require excessive maintenance. A number of other alternative methods of flow control are coming into use that will supplant the effluent flow control valve. For example, some plants have been constructed so as to split the flow nearly equally (influent flow splitting) to all the operating filters, usually by means of an influent weir box on each filter. A schematic diagram of such a filter is shown in Fig. 3. The advantages of this system include:

- Constant-rate filtration is achieved without rate controllers if the total plant flow remains constant.

- When a filter is taken out of service for backwashing or returned to service after backwashing, the water level gradually rises or lowers in the operating filters until

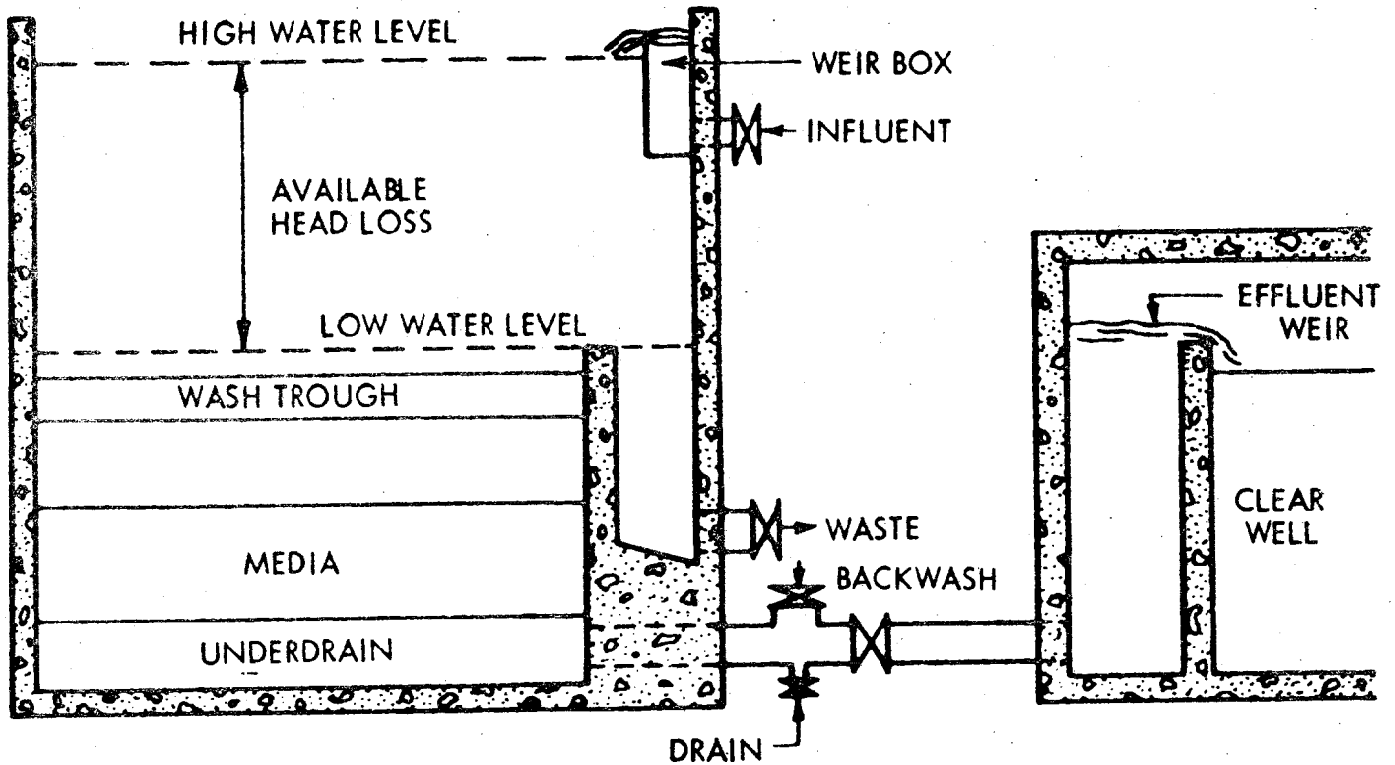


FIG. 3 Influent-control, constant rate filter

sufficient head is achieved to handle the flow. Thus, the rate changes are made slowly and smoothly without the abrupt effects associated with automatic or manual control equipment. This causes the least harmful effect to filtered water quality (5).

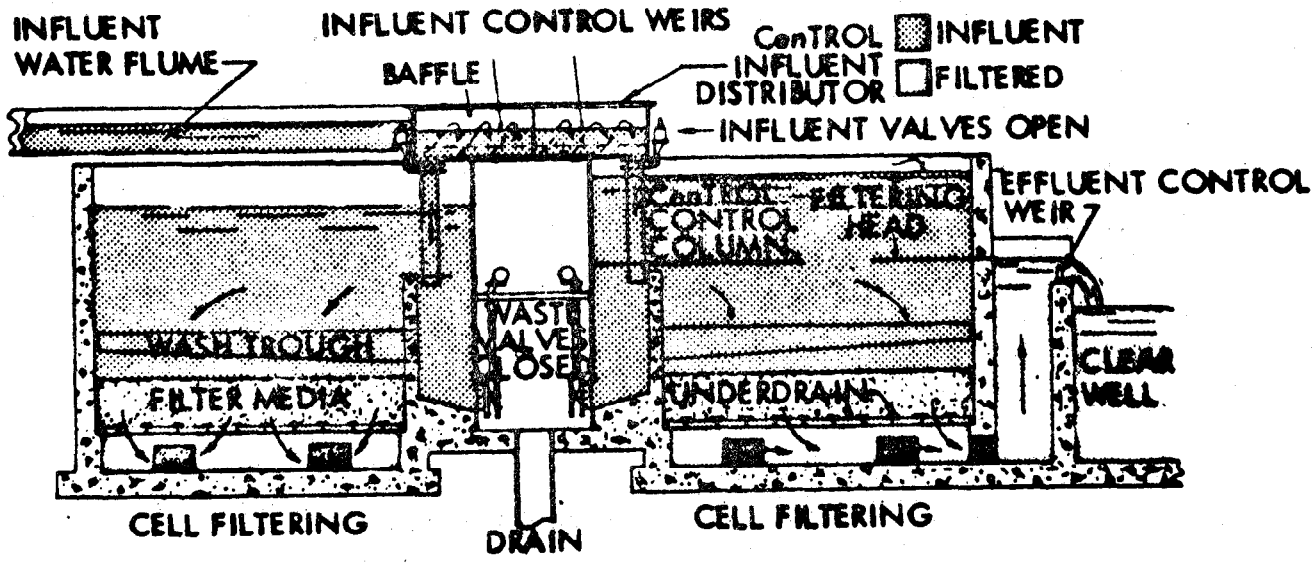
- The head loss for a particular filter is evident to the operator by the water level in the filter box andcan be read on a simple staff gage attached to the filter box wall, rather than by elaborate automatic remote reading devices. When the water reaches a desired maximum level (the desired head loss), backwashing of that filter is required. Effluent flow meters for each filter are not essential since the flow per filter is about the total plant flow divided by the number of filters in service.

- The effluent control weir must be located above the sand to prevent accidental dewatering of the filter bed. This arrangement eliminates completely the possibility of negative head in the filter and the well-known and undesirable problems which sometimes result from it.

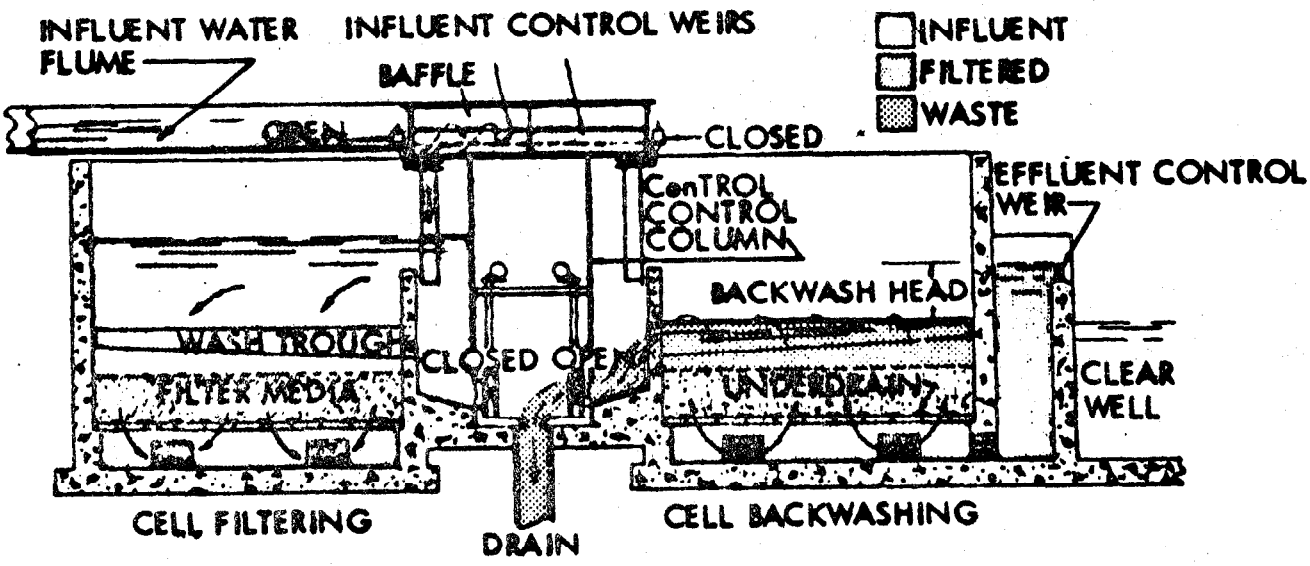
The only disadvantage of the influent-flow splitting system is the additional depth of filter box required. The depth of filter box must be increased 5 to 6 ft over conventional box depths to provide a reasonable available head loss. Figure 4 shows an influent-flow splitting filter in both filter and backwashing cycles.

It is also possible to operate a similar type of filter frequently referred to as declining-rate filtration. Declining-rate or variable declining-rate filtration has all of the advantages of constant-rate operation by influent-flow splitting plus some additional advantages. The system (based on an unequal and varying split of total flow to all operating filters) <u>should be considered in the design of all new wastewater filtration plants</u>. Despite its merits, it has not received enough explanation or attention.

The principal differences between declining-rate filtration and constant-rate filtration with influent-flow splitting is the location and type of influent arrangement. Figure 5 illustrates the desirable arrangement for new plants designed for variable declining-rate operation. The method of



FILTERING CYCLE



BACKWASH CYCLE

Fig 4 Operating cycles of influent-control, constant-rate filter

(132)

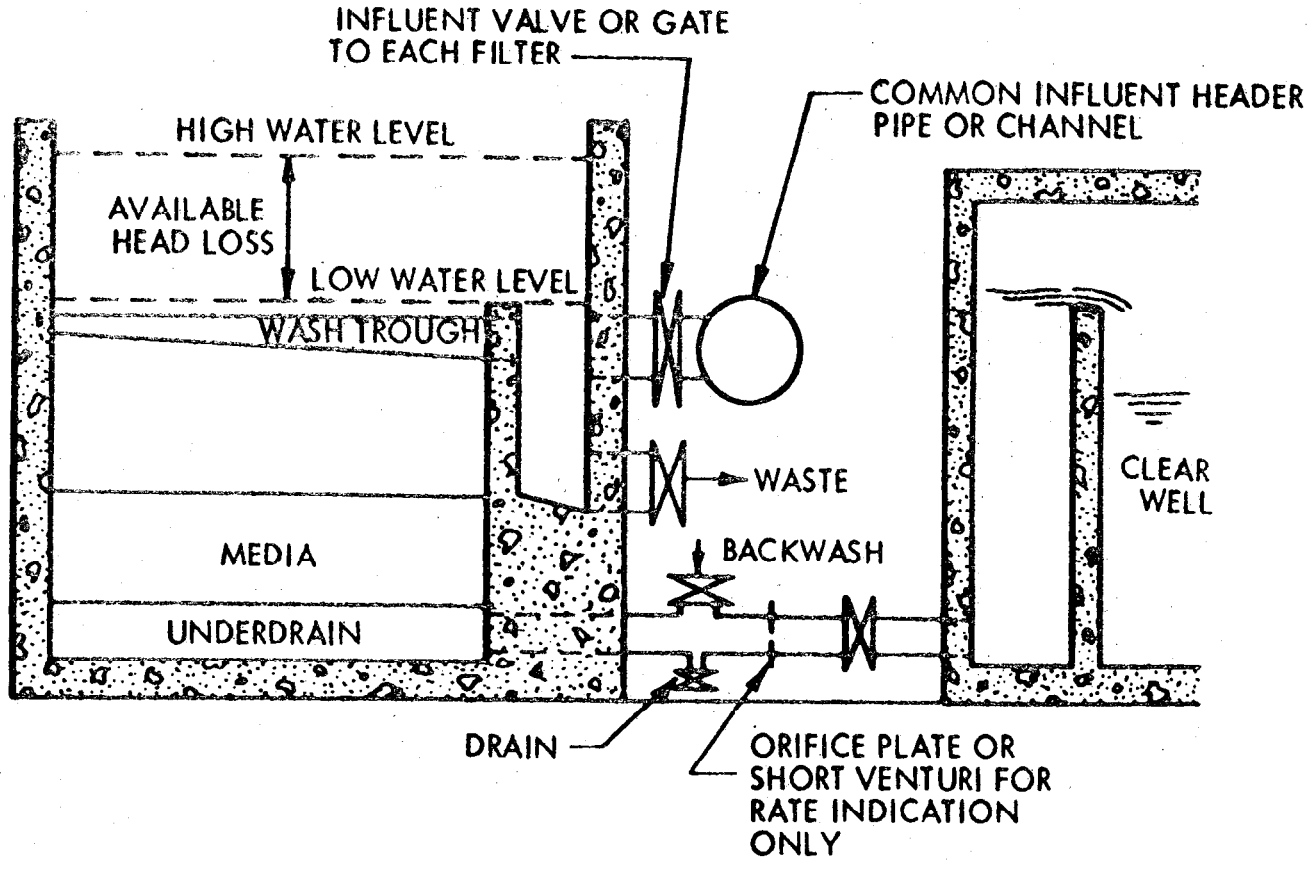


Fig 5 Declining-rate filter

operation is similar to that described for the filter in Fig. 3 with a few exceptions. The filter influent enters below the low water level of the filters through a relatively large influent header serving all the filters and a relatively large influent valve to each individual filter. Thus, the head losses in the header and influent valves are small and do not restrict the flow to a filter so the water level is essentially the same in all operating filters at all times.

As the filters served by a common influent header get dirty, the flow through the dirtiest filters decreases the most rapidly automatically causing the cleaner filters to pick up the capacity lost by the dirtier filters. The water level in all filters rises slightly as this happens and provides the additional head needed by the cleaner filters as they pick up the flow diverted to them by the dirtier filters.

This method of operation causes a gradually declining rate towards the end of a filter run. The advantages of declining rate operation over constant rate operation are (6, 7, 8):

- significantly better filter effluent quality than that obtained with constant-rate filter operation, and
- less available head loss needed compared with that required for constant-rate operation.

Filter effluent quality is adversely affected by abrupt increases in the rate of flow -- here the rate increases occur in the cleaner filters where they have the least affect on filter effluent quality (5). Rate changes throughout the day due to changes in total plant flow, both upward and downward, in all of the filters, dirty or clean, occur gradually and smoothly without any automatic control equipment.

The available head (or filter box depth) needed in operating declining-rate filter is less than that needed to operate a constant-rate filter for three reasons:

- The flow rate through the filter decreases toward the end of the filter run. The head loss in the underdrain and effluent piping system therefore decreases and becomes available to sustain the run for a longer period than would be possible under constant-rate operation with the same available head. The recovery of available head due to decreased flow at the end of the filter run is a function of the square of the flow rate. (If the head loss through the underdrain and effluent system were 4 ft at 4 gpm/sq ft, it would decrease to 2.2 at 3 gpm/sq ft [$4 \times (3/4)^2 = 2.2$] and provide an additional 1.8 ft of head.)

- As the dirty filters lose capacity, the load is redistributed among all of the filters. The resulting rate increase in each filter is very small and the water level on all filters rises very slowly until it approaches the maximum permissible water level. The filter with the longest time in service in backwashed, and when put back in service, that filter operates at the highest flow rate thereby causing the water level on all filters to decline to a new equilibrium level.

- The influent enters below low water level thereby saving the head on the weir and the free fall allowance needed where influent-flow splitting is used.

For the above reasons, declining rate filters are considered to be most applicable to the type of filter operation encountered in wastewater filtration.

3. APPLICABILITY TO THE SYSTEM IN QUESTION

In the wastewater treatment systems proposed for use in the Merrimack River basin, the granular filters would be used to reduce the suspended solids in the waste to a level which would be amenable to subsequent ion exchange treatment of the wastewater. In each treatment scheme, the filters would be located immediately following the expanded carbon beds used in adsorption of soluble BOD and in denitrification of the waste waters. In this location, the filters could be expected to operate effectively. The chief design problems would involve the following:

- The variable nature of the hydraulic rate of flow through the plant. What is the significant flow rate which should be used for design? Declining-rate filters -- at least without adequate operating data required to establish an operating model for determining optimum operational design -- should be designed for operating at a flow rate not to exceed 4 gpm/sq ft during the <u>significant total flow rate period</u>.
- The distribution of the total plant flow equally to all banks of filters which operate as a declining-rate unit.
- The choice of media sizes and depths and terminal pressure drops to be provided in all filters.

In the proposed flow scheme, declining-rate filters using automatic backwash with the following characteristics should provide the desired filtered water quality with run lengths of about 6-12 hours during the significant flow period:

Q = 4 gpm/sq. ft.

Media = 24 inches of 2-3mm Anthrafilt
20-24 inches of 0.75-0.90mm silica sand
or
18 inches of 2-3mm Anthrafilt
18 inches of 0.75-0.90mm silica sand
10 inches of 0.5mm garnet sand.

Terminal Head = 10 ft. of water

4. DESIGN CRITERIA WITH RANGES

Hydraulic design flow

In both the biological and physical-chemical flow scheme proposed for use in the Merrimack River basin, the mean daily flow from the study areas in 1990 have been estimated to be as follows:

Winnipesaukee ---------- 9.72 MGD

Lowell-Lawrence-Haverhill ---------- 121 MGD

Boston ---------- 500 MGD

Although it is important to know the average wastewater flow rate, consideration must also be given to the fluctuations in flow which occur over monthly, weekly, daily and hourly cycles and from dry to wet years. Table 1 shows typical (but not for the Merrimack study areas) monthly average flow rate variations expressed as a percentage of mean annual flow rate.

Table 1. Variation in monthly wastewater flow rates expressed as percentages of the average yearly flow.[3]

Month	Total
January	91
February	90
March	105
April	112
May	121
June	111
July	102
August	99
September	87
October	104
November	91
December	87

The variation in daily flow rate is shown in Table 2. Table 3 shows how the hourly flow varies within any one day with maximum hourly flow occurring around 10 AM to 2 PM (Figure 6), probably the significant hydraulic period.

Table 2. Variation in daily flow rate expressed as percentages of the average weekly flow rate.

Week day	Total
Sunday	82
Monday	106
Tuesday	104
Wednesday	103
Thursday	103
Friday	104
Saturday	98

Table 3. Variation in hourly flow rate expressed as percentages of the average daily flow.

Hour of day	Percentage	
8 AM	52	
10 AM	142	4-hr pk. = 144
11 AM	150	
12 Noon	145	
2 PM	140	
4 PM	125	
6 PM	115	
8 PM	120	
10 PM	108	
12 Midnight	103	
2 AM	80	
4 AM	50	
6 AM	42	

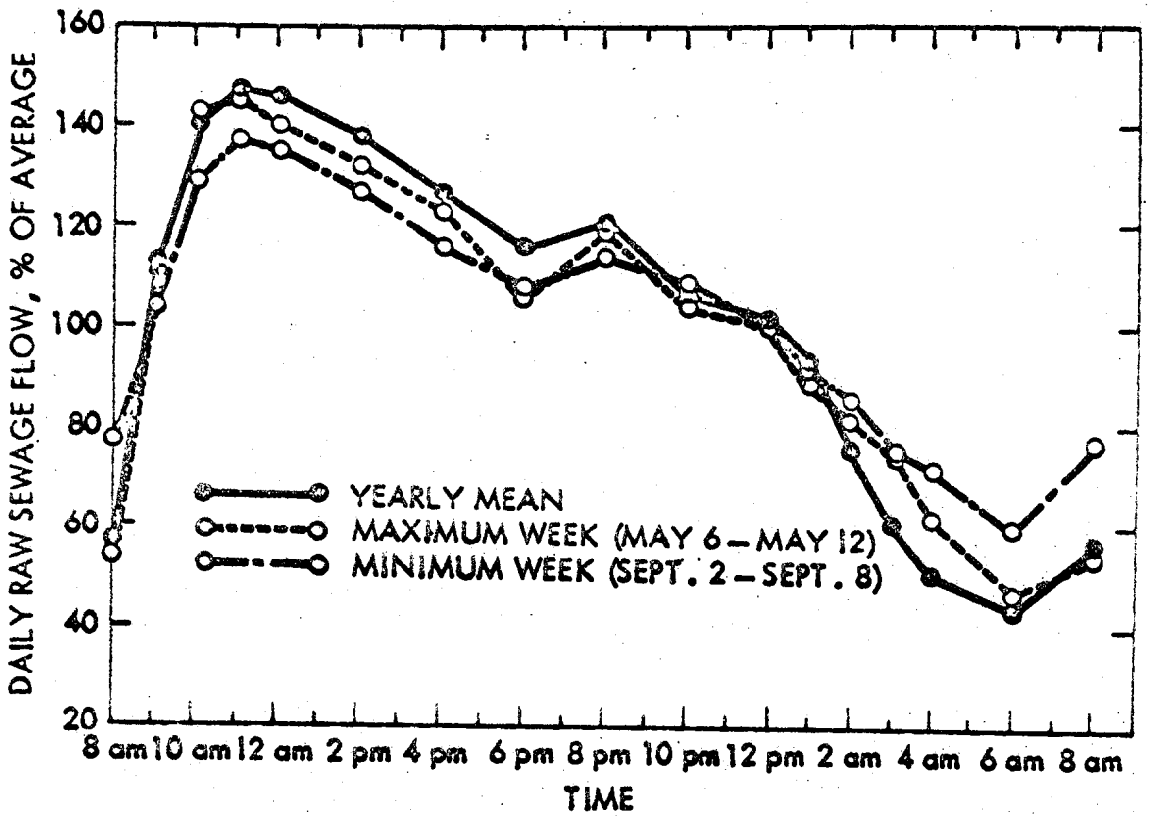


Figure 6. Ranges in typical 24-hr wastewater flow pattern.

In general, patterns of variation in flow rate are relatively consistent as long as there are no major changes in population characteristics. The above flow patterns (dry weather flows) do not include the excessive flows due to combined sewers.

The maximum hourly wastewater flow reaching a water pollution control plant would be expected to occur on the maximum day of the month of highest flow during a wet year. This maximum flow would be represented by the product of the average yearly flow rate and the extreme variation in yearly, monthly, daily, and hourly flow rate expressed as a decimal fraction. Then

$$Q_p = \bar{Q} \times \frac{M_p}{\bar{Y}} \times \frac{W_p}{\bar{M}} \times \frac{D_p}{\bar{W}} \times \frac{H_p}{\bar{D}} \times \frac{\bar{Y}_w}{\bar{N}} \qquad (1)$$

where:

Q_p = peak hourly flow, mgd

$\bar{Q}$ = average yearly flow rate, mgd (500 mgd for Boston)

$\frac{M_p}{\bar{Y}}$ = ratio monthly peak flow rate to average yearly flow rate (Table 1, May)

$\frac{W_p}{\bar{M}}$ = ratio weekly peak flow rate to average monthly flow rate (Table 2)

$\frac{D_p}{\bar{W}}$ = ratio daily peak flow rate to average weekly flow rate (Table 2, Monday)

$\frac{H_p}{\bar{D}}$ = ratio hourly peak flow rate to average daily flow rate (Table 3, 4-hr peak)

$\frac{Y_w}{\bar{N}}$ = ratio average yearly flow rate in wet years to average yearly rate for normal years.

Normally, we should not design filters (or any other part of the plant) for the single largest hourly flow rate possible since this flow rate may occur for only a few hours in any one year. In normal operation of a treatment plant, the peak flow covers several hours and can be smoothed out by using the incoming sewer line and pumping station storage volume for flow equilization. Consequently, the peak flow rate for a 4-hr period should be satisfactory for peak hydraulic and organic load design purposes.

The relationship between the average and peak 4-hr wastewater flow rate for the city represented by the data included herein during wet years would then be

$$Q_p = \bar{Q} \times 1.21 \times 1.00 \times 1.06 \times 1.44 \times 1.11$$

$$Q_p = \bar{Q} \times 2.05.$$

The actual peak flow rate which should be used for design purposes must be selected on the basis of the ultimate treatment needs. If short periods of hydraulic overload and reduced treatment efficiency can be accepted, the design flow can be reduced, thereby reducing the size of the treatment units. If such periods of overload cannot be accepted the plant might have to be designed for the most extreme load expected within the design life.

With the flow schemes proposed, data should be accumulated so that the filter design can be based on the plant flow during the critical 4- or 6-hour period.

For the three study sections a very, roughly, approximate, filter design flow was arrived at as follows:

	Current Sewered Population	Flow, MGD	Flow, gpcd	1990 Population	Q peak** / Q mean	FILTER*** DESIGN Q, MGD
Winnipesaukee	50,500(?)	8.0	158	61,500	2.51	18.3
L-L-H	257,000	70.0	272	446,000	1.81	164
Boston	1,880,000	424.0	225	2,220,000	1.38	690****

* BOD = 8,600 #/day, 8,600 #/0.17 # = 50,500 population
** Q peak/Q mean = $5/P^{0.167}$ (9)
*** Filter design Q = 1990 MeanxQ peak/Q mean x 0.75
0.75 arbitrarily used to convert from peak instantaneous flow to peak 6-hr flow
**** Factor 0.75 neglected

Thus, the critical flows for FILTER DESIGN became:

	MGD
Winnipesaukee	18.3
L-L-H	164.
Boston	690.

Critical filter sizes

In order to be able to distribute raw water and collect backwash water effectively, the maximum size of any one filter should be limited to about 850 sq. ft. This would limit the water filtered by any one filter to a maximum of

4 gpm/sq ft x 850 sq ft x 1,440 min/day = 5.90 MGD

In normal practice, 4 filters are usually included in a single declining rate unit and thus a single unit would have a maximum capacity of 19.6 MGD (4 x 4.90 MGD).

In plant operation, it is not unusual to have 1 filter or even 1 unit of filters "down" for service at any one time. If one filter is down for service the effective flow rate on the other 3 filters would be 5-1/2 gpm/sq. ft. If an entire battery of filters were out of service, it would be necessary to have a minimum of 3 declining-rate filter units (or a minimum of 12 filters) in any one wastewater filtration plant to provide a maximum filtration rate of 6 gpm/sq. ft. when all remaining filters were in service. When one of these filters was being backwashed, the average filtration rate on the other filters would be 6.85 gpm/sq. ft. Thus, critical design criteria would be:

Maximum Filter size = 850 ft^2

Minimum No. of declining-rate units = 3 (absolute)
= 4 (preferred)

Using these criteria, the filter design requirements for each of the study areas would be as in Table 4.

Table 4. Typical filtration plant characteristics for study areas.

	Winnipesaukee		L-L-H		Boston	
	Mean Flow	Design Flow	Mean Flow	Design Flow	Mean Flow	Design Flow
Total Capacity, mgd	9.72	18.3	121	164	500	690
No. of Units, (4 filters each)	3	3	7	9	27	37
Flow per Unit, mgd	3.2	6.1	17.3	18.2	18.5	18.7

Filter layout and details

For maximum construction economy, the 4-filter (or 6-filter) declining-rate units should be rectangular in plan so as to promote common wall construction. In the filtration plant layout, factors to be considered in design include:

- Influent distribution uniformly to all filter units is a critical item in plant design. Where only 3 to 6 units are involved, an overflow-weir type splitter box might be appropriate. Where more units are involved, an influent flow channel might be used alone one side of the units with parshall-flume control to each unit. Equal flow distribution is essential.

- An adequate supply of backwash water (15-25 gpm/sq ft) is essential for proper backwashing. For plants using 3-6 units, no separate backwash pumps need be used if construction typical of Fig. 4 is used. However, all filter units must be connected to the same overflow weir so filters in all units can be used to supply backwash water. (Three filters can only supply 12 gpm/sq ft of backwash water to the unit undergoing backwash.) To obtain 15-25 gpm/sq ft, additional water is needed from other filtering units. With more than 3-6 units, the filter walls can be reduced in height to reduce total plant cost by using separate backwash pumps and a filter section as shown in Fig. 3.

- The height of the filtering cell walls would be approximately as follows:

	Without Backwash Pumps	With Backwash Pumps
Head Loss to Weir Level	10 ft.	10 ft.
Weir Level to Wash Trough	4 ft.	0 ft.
Wash Trough to Media	4 ft.	4 ft.
Media	4 ft.	4 ft.
Underdrain	1 ft.	1 ft.
TOTAL	23 ft.	19 ft.

- The filter design should provide for effective air-water backwash in sequence as follows:

  - Air wash of 4-5CFM/sq. ft. for 2 minutes.

  - Air-water wash for 2 minutes.
    Air at 4-5 CMF/sq. ft.
    Water increasing to 15 gpm/sq. ft.

  - Water wash of 15-20 gpm/sq. ft. for 10 minutes.

- The total backwash cycle of 30 minutes would be distributed as follows:

  - Shut off filter, flow time to drain water to about media level ---------- 15 minutes
  - Air-wash ---------- 2 minutes
  - Air-Water wash ---------- 2 minutes
  - Water wash ---------- 10 minutes
  - Return to service ---------- 1 minute

  TOTAL 30 minutes

5. PRETREATMENTS REQUIRED

In the flow schemes included in this study, no special additional pretreatment is required prior to tertiary filtration of the wastewater. The use of additives such as polyelectrolytes is not anticipated, but the filtration plant should be designed to provide for inclusion of such facilities (storage, mixing, feeding) if operating experience requires them.

6. EFFLUENT QUALITIES

The operation of the proposed dual media filters at a rate of 4 gpm/sq. ft. to a terminal head loss of 10 ft. of water at the plant flow rate during the 6-hour significant period should give 12 hour filter runs and an effluent quality not exceeding:

BOD_5, mg/l	2
SS, mg/l	2 (4-5 Turbidity Units)
COD, mg/l	5
N, mg/l	1
P, mg/l	<1
TDS, mg/l	275

7. PROCESS EXPERIENCE TO DATE

Filters are normally designed for the purpose of removing suspended solids, -- and only suspended solids. Thus, filtration applications which are designed for the removal of soluble BOD, for example, are bound to fail. Filters have been and can be applied successfully in tertiary wastewater treatment plants for the following purposes:

- Reduction of SS and BOD from trickling filter and/or activated sludge plants following final sedimentation. Plant efficiency in BOD removal is a function of what proportion of the BOD is represented in the SS.
    - Plants operate w or w/o chemical pretreatment.
- Reduction of SS, their associated BOD, and residual chemical precipitant from trickling filter and/or activated sludge plants <u>or from primary</u> effluents treated with iron or aluminum salts or lime for the purpose of improving solids removal in subsequent sedimentation <u>and</u> for precipitation of phosphates from the plant discharge.
- Since expanded carbon beds can operate effectively with some suspended solids, filters can be used to remove residual biological suspended solids, residual chemical precipitants, and degraded carbon particles from the effluents of expanded carbon beds.

In practice, ample evidence is available to demonstrate that wastewater effluents can be clarified effectively and dependably. It is not difficult to obtain a filtrate that will have a turbidity of less than 5 turbidity units, the maximum turbidity permitted in drinking water according to the U.S. Public Health Service drinking water standards. The difficulty with tertiary filtration plants lies in being able to obtain reasonable, economic filter run lengths which will permit operation with a reasonable consistancy under conditions of both flow variation and waste strength variation. Filters cannot be designed for average flow conditions and be expected to perform adequately under critical flow and solids loading conditions.

Supposedly, we should be able to learn from the experience of others. To summarize the experience of others, Table 5 was prepared as a summary of the published experience with Granular Filters for Wastewater Filtration. This is not a complete list, since there should be a significant number of FWQA demonstration project reports with more complete operating data for more recent American experience. It is safe to conclude, however, that European experience with and use of filtration for improving wastewater quality is more extensive than American experience.

Table 6 is a separate list of the references cited in Table 5 kept separate for use as a ready list of wastewater filtration experience sources.

In view of the extensive published experiences summarized in Table 5, we might expect that current tertiary filtration experience would be eminently successful. Unfortunately, it is not. To the authors personal knowledge, there are no tertiary filtration plants yet built in Iowa and none in any significant numbers in any surrounding states except in Illinois. Illinois plants have been designed in accordance with the (Illinois) Sanitary Water Boards Technical Release 20-24 which requires that wastewater discharged to a stream with less than 1:1 dilution have a BOD less than 4 and suspended solids less than 5 mg/l. To approach these levels, Illinois requires filtration of the effluent prior to discharge as the first stage treatment in a program of improving stream water quality. As of about May, 1970, for example, the 5-county area (Cook-DuPage-Kankakee-Lake, and Will Counties), near Chicago had approved 11 such tertiary filtration plants, about 1/2 were in operation, none of them really successfully. The meager data on these can be summarized in Table 7.

Table 5.

SUMMARY, GRANULAR FILTERS FOR WASTEWATER FILTRATION

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

Table 5 (cont'd.)

SUMMARY, GRANULAR FILTER FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Author's Remarks
South River plant, N. J. (from Streander[19], 1940)	Primary settled effluent (No chem.)	Sand 1.0 U.C. 1.5	6	0.64		1.9	132	70	47			28	37% treated water used	Total washing time/day = 190 min.
Sayerville, N. J. (from Streander[19], 1940)	Primary settled effluent (No chem.)	Sand 1.0 U.C. 1.5	6	0.6		2.7	154	76	51			13	47% treated water used	Total washing time/day = 135 min.
Sayerville, N. J. (from Streander[19], 1940)	Primary settled effluent with chem.	Sand 1.0 U.C. 1.5	6	0.6		3	40	20	50			9	41% treated with water	Total washing time/day = 120 min.
Laughlin filter, Atlantic City (from Streander[19], 1940)	Primary settled effluent	Sand, E.S. 0.4 ~ 0.5 U.C. 1.8 ~ 2.0	6	2			88	39	56	187	137	26		Reduction in BOD are not proportional to the reduction of SS
Aucor Sewage Works, pilot plant, South Africa (from Vosloo[24], 1947).	from humus tank effluent	Sand 0.5 ~ 1.7	29	2	10	6~18	24.3	0.7	97				5% of total flow used for backwash	
Pilot plant at Luton, England 1949-1950 (from Pellet, Callett, Woddington[17], 1952)	humus tank effluent	Sand 0.85~1.7	24	2	5.6	24	24	3.4	86	26	10	61.5	2-3% of treated water, 13.3 gpm/ft^2 rate	No appreciable difference in head loss for sand and filters with 2 ft and 3 ft 6 in depth, respectively. Also little difference in effluents from two filters at rates below 2.92 gpm/ft^2, but at higher rates, the effluent from filters of 3'-6" of sand was slightly superior.
	humus tank effluent	Anthracite 1~2	24	2	7	24	25	3.3	87	26	9	65.5	2~3% of treated water, 10 gpm/ft^2 rate	
	humus tank effluent	Sand 0.85~2	24	2.34~3.42			19.3	0.9	95	23.8	7.2	70		
		"	42	2.34~3.42			19.3	0.7	96	23.8	6.3	74		

Table 5 (cont'd.)

SUMMARY, GRANULAR FILTER FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Author's Remarks
Luton Plant, England May 1951 ~ April 1952 (from Evans, Roberts[5], 1952)	from humus tank effluent	Sand 0.85~1.67	8	3.33	8~9	24	15.4	4.2	73	13.0	6.7	49	By air-scoar, 2.5% of treated water	1. During storm flow and high loading, backwash every 12 hrs. 2. Lower DO in filter efflents. 3. Oxidation of ammonia range from 35 to 68%. 4. Cost per lb of SS removed = 4.25 d Cost per lb of BOD removed = 7 d. 5. Visible color reduce 6.5%; thiocyanate, 50%; phenols, 8%.
Full Scale Sand Filter at Luton, England (1951 ~ 1954)(from Evans, Roberts[6], 1955)	Humus tank effluent	Sand 0.9 ~ 1.67		3.33		12	16.1	4.6	72	12.4	6.2	50		SS effluent was slightly higher during winter conditions than summer conditions.
Full scale sand filter at Luton, England (From Evans[4], 1957).				2.5		24	13.4	3.3	76	9.2	4.0	56.5	Use 3% of treated water	
Los Angeles Hyperion treatment plant (preliminary tests) (from Laverty et al.[11], 1961).	Activated sludge effluent	Sand E.S. 0.95 U.C. 1.6	11	2			26.1	12.7	46	9.8	4.8	51	automatic backwash	
Pretoria sewage treatment works Johannesburg, South Africa (from Bolton & Klein[1], 1962; Keefer[10] 1962)	Secondary effluent	Sand 20/30 mesh (British)	21	2.8	6.5	6~10	22.0	4.6	79				95% of treatment water	

Table 5 (cont'd.)

SUMMARY, GRANULAR FILTER FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Author's Remarks
Peoria Sanitary District Treatment Works (Pilot) (from Fall & Kraus', 1964).	high rate activated sludge effluent			1.12			1.60	8	50	28.3	17	40		
Research at Iowa State University. Pilot plant was located at Ames Sewage treatment plant. (Merry[4], 1965)	Trickling filter final effluent	Sand		2		10	19.8	5.7	71.2	56	24.2	56.8	20-25 gpm/ft^2 for 10 min.	
		ES 0.55	24	4	6.25	6.25	18.5	6.0	67.6	52.6	23.4	55.5		
		UC 2.36		6		4.67	17.7	6.3	64.4	50.4	24.4	51.6		
Philowath Municipal sewage treatment plant, Oregon (from Culp, Hansen[2], 1967)	Extended aeration effluent						59	4	93	26	2.5	90		Remove phosphate by adding chemical
Gravity sand filter and pilot up-flow filter at Luton, England 1966 (from Naylor, Evans, and Dunscombe[15], 1967)	final settled down-flow full scale	Sand 0.85~1.7	36	4.2	2	12	12.8	7.5	41	5.3	3.1	42		up-flow filter was superior in every respect
	up-flow pilot plant	Sand 0.85~1.7	60	4		8	12.8	5.7	56	5.3	2.6	51		
Cambridge, England, 1967 (by Holding Discussion, 1967)	final settled effluent down-flow	Sand 0.85~1.7	24	2.5	3	6	20-40		77					1. Tertiary treatment could only be successful when the influent was well oxidized
	up-flow	Sand 0.85~1.7	24	2.5		24	20-40							2. Effluent poorer at higher rate

Table 5 (cont'd.)

SUMMARY, GRANULAR FILTER FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max. terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Author's Remark
West Hertfordshire authority, England, 1968 (from Wood, Smith, Murry[23] 1968)	up-flow filter final effluent of trickling filter	Sand 1~2	60	1.8	5		44.3	1.9	95.7	57.7	3.9	93.1	Air scour then backwash with untreated effluent	1. Sharp fall in efficiency for flow greater than 4 gpm/ft^2 2. Little or no nitrification or denitrification took place in the filter
		Sand 1~2	60	3.33	5		37.3	3.7	90.1	53	4.6	91.3		
		Sand 1~2	60	4.16	5		55.5	7.1	87.1	42	5.6	86.6		
		Sand 1~2	60	5.0	5		37.3	9.9	73.5	34.6	14.7	57.5		
Letchworth Plant, England 1966 ~ 1967 (from Truesdale and Birbeck[21,22] 1968)	final effluent of activated sludge up-flow filter pilot plant	Sand 1~2	60	4.4			17.2	6.9	60	19.2	9.0	53	Air scour then washed by upward flow to about 11.7 gpm/ft^2. 6% treated water used	Effluent from filter was inferior at the higher rates. (>6.6 gpm/ft^2)
Hanover treatment plant Metro Chicago, Harding design filter (from Sosewitz & Bacon[18] 1968 from Lynam, Ettelt, & McAloon[13] from Lynam, Bacon[12], 1970)	final effluent of activated sludge sludge	Sand ES 0.58 UC 1.62	12			4~5			77			89	traveling backwash	1. Efficiency of tertiary treatment depend on the efficiency of secondary treatment. 2. Additional solids removal provided by coagulation with alum plus polymer was not sufficient to warrant inclusion with filtration into tertiary treatment design.
Pilot scale Sand filter at Derby, England (from Oakley & Cripps[16], 1969-Joslin & Greene[9], 1970)	Trickling filter final effluent	comm. Sand 1.2~2.4	24	2.5	1	14-17			59-60			23-38		
				3.33	1	20			57-68			50-64		
				5.0	1	17-18			47-58			32-37		
		comm. Sand 1.2~2.4	36	3.33	1	19			65			67		
				5.0	1	16-17			60			43-46		
		Sand 1.2~1.68	24	2.5	1	10-12			59-60			54		
				3.33	1	11-15			52-64			53-65		

Table 5 (cont'd.)

SUMMARY, GRANULAR FILTER FOR WASTEWATER FILTRATION

Source Information	Waste-water Charact.	Type, Size Filter media (mm)	Depth (in)	Rate (gpm/ft^2)	Max terminal head (ft water)	Run length (hr.)	SS (ppm) Infl.	SS (ppm) Effl.	SS (ppm) % Removed	BOD_5 (mg/l) Infl.	BOD_5 (mg/l) Effl.	BOD_5 (mg/l) % Removed	Backwash Information	Author's Remark
Pilot scale Sand filter at Derby, England (from Oakley & Cripps[16], 1969-Joslin & Greene[9], 1970)	Trickling filter final effluent	Three-layer filter 8", 7/12 Anthracite		2.5	1	13			64			69		
		8", 12/14 Sand		3.33	1	13-17			57-70			59-71		
		8", 18/22 Garnet		5.0	1	12-14			53-65			34-46		
		up-flow 4', 1~2. Sand		3.33	1	25-26			57-74			60-67		
		2', 2" Gravel		5.0	1	12-17			47-58			48-50		
High Rate filtration of Mill Scale waste (from Donovan[3] 1970)	Mill scale waste	Anthracite 5.1	84	8	15	46	150	11	92.5				With the relatively coarse media, only slight expansion, along with the air, is required to break up any agglomerates. Air scour for 5 min. at 8 cfm/sf then flush with water at 30 gpm/sf for 10 min.	The cationic polyelectrolyte produced the best results. The cost of feeding polyelectrolyte was less than $1.00 per million gallon water filtered.
				16	15	38	150	36	76					
				23	15	33	150	83	45					
				30	15	23	150	85	43					
High Rate filtration of Mill Scale waste (from Donovan[3] 1970)	Mill scale waste	Sand 2-3	84	8	15	20	150	2	99					
				16	15	9.5	150	4	98					
				23	15	20.5	150	30	80					
				30	15	3.5	150	7	95					
	Mill scale waste	3'-2.3 Sand 4'-5.1 Anth.		8	15	28	150	3	98					
				16	15	20.5	150	3	98					
				23	15	12	150	10	93					
				30	15	8.5	150	9	94					
Research at Stanford Univ. (from Tchobanoglous & Eliassen[20] 1970)	settled effluent Bloc pilot scale activated sludge unit													Rational equations were developed to predict filter performance.

Table 6.

REFERENCES

1. Bolton, R. L. and Klein, L., "Better Sewage Effluents: thier Need and Attainment." Public Works, 93, No. 10: 109-112. 1962.
2. Culp, G. L. and Hansen, S. P., "Extended Aeration Effluent Polishing by Mixing Media Filtration." Water and Sewage Works, 114: 46. 1967.
3. Donovan, E. J. Jr., "High-Rate Filtration of Industrial Wastes," Water Quality Improvement by Physical and Chemical Processes, Water Resources Sumposium No. 3, University of Texas Press, 167-189. 1970.
4. Erans, S. C., "Ten Years of Operation and Development at Luton Sewage Treatment Works," Water and Sewage Works, 104, No. 5:214-219. 1957.
5. Erans, S. C. and Roberts, F. W., "Twelve Months' Operation of Sand Filtration and Micro-Straining Plant at Luton," J. Proc. Inst. Sew. Purif., 333. 1952.
6. Evans, S. C. and Roberts, F. W., "Recent Developments in Sewage Treatment at Luton," J. Proc. Inst. Sew. Purif., 225. 1955.
7. Fall, E. B. Jr., and Kraus, L. S., "Tertiary Treatment for High Rate Activated Sludge Effluent," Unpublished paper presented at the 37th Central States Water Pollution Control Association Meeting. June, 1964.
8. Holding, J. C., Discussion on paper by Truesdale, G. A., "Tertiary Treatment Processes for Sewage Works Effluents," J. Inst. Water Poll. Control, 66, 380. 1967.
9. Joslin, J. R. and Greene, G., "Sand Filter Experiments at Derby," J. Inst. Water Poll. Control, 611. 1970.

Table 6 (cont'd.)

10. Keefer, C. E., "Tertiary Sewage Treatment," Public Works, 93, No. 11: 109-112. 1962.

11. Laverty, F. B., Meyerson, L. A. and Stone, R., "Reclaiming Hyperion Effluent," American Society of Civil Engineers Sanitary Engineering Division Proceedings 87, No. SA6, Part 1: 1-40. 1961.

12. Lynam, B. T. and Bacon, V. W., "Filtration and Micro-Straining of Secondary Effluent," Water Quality Improvement by Physical and Chemical Processes, Water Resources Symposium, No. 3, University of Texas, 132-148. 1970.

13. Lynam, B. T., Ettelt, G. and McAloon, T., "Tertiary Treatment at Metro Chicago by Means of Rapid Sand Filtration and Microstrainers," JWPCF, 41: 247-279. 1969.

14. Merry, K. J., "Tertiary Treatment of Domestic Waste Water by Rapid Sand Filtration," Unpublished MS Thesis, Iowa State University, Ames, Iowa 1965.

15. Naylor, A. E., Evans, S. C. and Dunscombe, K. M., "Recent Developments on the Rapid Sand Filter at Luton," J. Inst. Water Poll. Control, 66, 309. 1967.

16. Oakley, H. R. and Cripps, T., "British Practice in the Tertiary Treatment of Wastewater," JWPCF, 41: 36-50. 1969.

17. Pettet, A. E. J., Collett, W. F. and Waddington, J. I., "Rapid Filtration of Sewage Effluents through Sand and Anthracite," Sewage and Industrials Wastes, 836, 1952.

18. Sosewitz, B. and Bacon, V. W., "Chicago First Tertiary Treatment Plant," Water and Wastes Engineering, 5, 9: 52. 1968.

Table 6 (cont'd.)

19. Streander, P. B., "Sewage Filtration With Silica Sand Filter," Part I - Water Works and Sewage, 351, Aug. 1940; Part 2 - Water Works and Sewage, 422, Sept. 1940; Part 3 - Water Works and Sewage, 493, Oct. 1940.

20. Tchobanoglous, G. and Eliassen, R., "Filtration of Treated Sewage Effluent," JSED, ASCE, Vol. 96, No. SA2, April 1970.

21. Truesdale, G. A. and Birkbeck, A. E., "Tertiary Treatment Processes for Sewage Works Effluents," Water Pollution Control, 66, 371. 1967.

22. Truesdale, G. A. and Birkbeck, A. E., "Tertiary Treatment of Activated Sludge Effluents," Water Pollution Control, 67, 483. 1968.

23. Wood, R., Smith, W. S. and Murray, J. K., "An Investigation into Upward Flow Filtration," Water Pollution Control, 67, 421. 1968.

24. Vosloo, P. B. B., "Some Experiments on Rapid Sand Filtration of Sewage Works Effluent Without Coagulation," J. Proc. Inst. Sew. Purif., 204. 1947.

25. Zack, S. I., "Mechanical Filtration of Effluents," Sewage Works Journal, Vol. 9, No. 3, 466. 1937.

Table 7. Tertiary filtration plant locations.

	Plant Location	Filters	Biological Plant	Remarks
1.	Addison South*	S and L, Graver 3, 10 or 12-ft. Monovalve 5GPM/sq. ft., Air wash-4 CFM/sq. ft., 5 minutes, 24-in. sand, 24-in. anthrafilt, 4800 gals/backwash, 6-8 hour runs.	S and L Oxigest Trickling Filter	Underdrains plugged 2.1MGD Cost - $40,000
2.	Addison North*	S and L, Graver 3, 8-ft. Monovalve	S and L Oxigest	Opened July 1970. Filter design for 3x Mean Flow.
3.	Bensenville	Neptune Micro-Floc	-----	------
4.	Park Forest South	Walker Process 4-gravity filters w/air wash	-----	------
5.	Lisle (1971 completion)	Eimco 6, 12-ft. diameter, dual valve, air wash	Extended aeration	Multi-media
6.	Lake Zurick	Eimco 4, 8-ft. filters, air-wash, 5GPM/sq. ft.	Walker Contact-Stabilization plant	Very short runs. Additional filters to be added to reduce rate.
7.	Barrington	Neptune, Micro-Floc	-----	------
8.	Hanover Park (M.S.D.) (experimental)	Walker Process	Follow Dorr Clarifier-flocculators	Experience Published (4)
9.	Woodale South	S and L, Graver	-----	------
10.	Glendale Heights (contract awarded)	S and L, Graver	-----	------
11.	Lake County Public Works Dept. (not built)	Eimco	-----	------

* Su[illegible] Anthony LaRocca

(Information about these plants can be obtained from Carl Blomgren, Engineer in the Chicago Office of the Illinois Sanitary Water Board.)

In early 1970, Anthony LaRocca, Supt. of the Addison plants gave an oral report at an operators meeting indicating the type of operation secured at the Addison South plant. SS applied to the filters averaged 45-50 mg/l and varied from 35-90 mg/l. With applied SS of 35 mg/l, the effluent carried from 2-10 mg/l. With applied SS of 90 mg/l, the effluent contained from 14-27 mg/l. With applied BOD of 44 mg/l, the effluent was at 20 mg/l.

Other tertiary treatment plants known to be in operation include:

1. General Filter Company (Ames, Iowa) plant at:

   Tree Top (General) Apartments
   Atlanta, Georgia

   This is a 98 gpm dual valve-dual media fully automatic plant with air wash operating on an apartment complex waste treated using an extended aeration package plant. Plant performance reported to be excellent.

2. Walker Process Plant at Atlanta, Georgia

   Dual-media gravity filters for filtration of extended aeration effluent. Unsatisfactory performance, since heavy loads bypass plant into chlorine-contact tank from which filter backwash is drawn. Media-retaining strainers clog with backwash SS.

3. General Filter Company

   C. D. Hulten Enterprises
   Woodbridge, Virginia

   A tertiary filtration plant at Dale City, Virginia following lime precipitation of phosphates following biological secondary treatment. Uses GFC Contra-Flo's and CenTrol filters for treatment rate of 2100 gpm (3 MGD). Plant privately owned. Plant to start in operation almost immediately.

4. Eimco Corporation
St. Lake City, Utah

Under WQO, EPA Contract WP-14-12585, Eimco is operating a 100 gpm pilot plant for physical-chemical treatment including use of powdered activated carbon. Tertiary filters (0.9mm coal, 0.5mm sand at 2-3gpm/ft^2) using dual media are used. Backwash with air at 4-5 CFM/sq. ft. for 1 minute, air-water backwash for 1/2 minute, and water backwash at 25 gpm/sq. ft. for 5 to 8 minutes. Results reported to be excellent.

8. CAPITAL AND OPERATING COSTS

The costs involved in tertiary filtration of wastewaters is even more difficult to obtain than operating experience data. Usually, filtration cost data are buried in total general contract bidding and are not too available. However, an attempt was made to modify the typical cost model for filtration of industrial waste supplied with the preliminary information to the consultants. Figure 7 shows the capital costs for tertiary treatment filters derived from several sources:

- Cost of Clean Water, Vol. IV, Cyrus Wm. Rice and Company, 1968.
- Weber et al, JWPCF, 42:83, January, 1970.
- Estimate by Baumann based on preliminary plant design characteristics (Table 4) and 1969 equipment cost data and concrete cost of $300/cu. yd.

The costs of the mean flow and design flow plants for the three study areas were evaluated using the techniques described in a recent paper (2). Appendix A lists the estimated unit costs for equipment for each of the 6 plants. Appendix B lists the costs for concrete for the six plants involved for concrete-in-place costs of $200 and $300 per cu. yd. Table 8 lists the capital costs of the various tertiary filtration plants for the study areas. These cost data indicate that very little unit cost savings is effected as plant size increases above about 60-80MGD, since the larger size plants merely require a much larger number of the maximum size filter unit. Some savings can, of course, be realized by common wall construction, etc., but Fig. 7 should provide an adequate model for determining preliminary capital cost data.

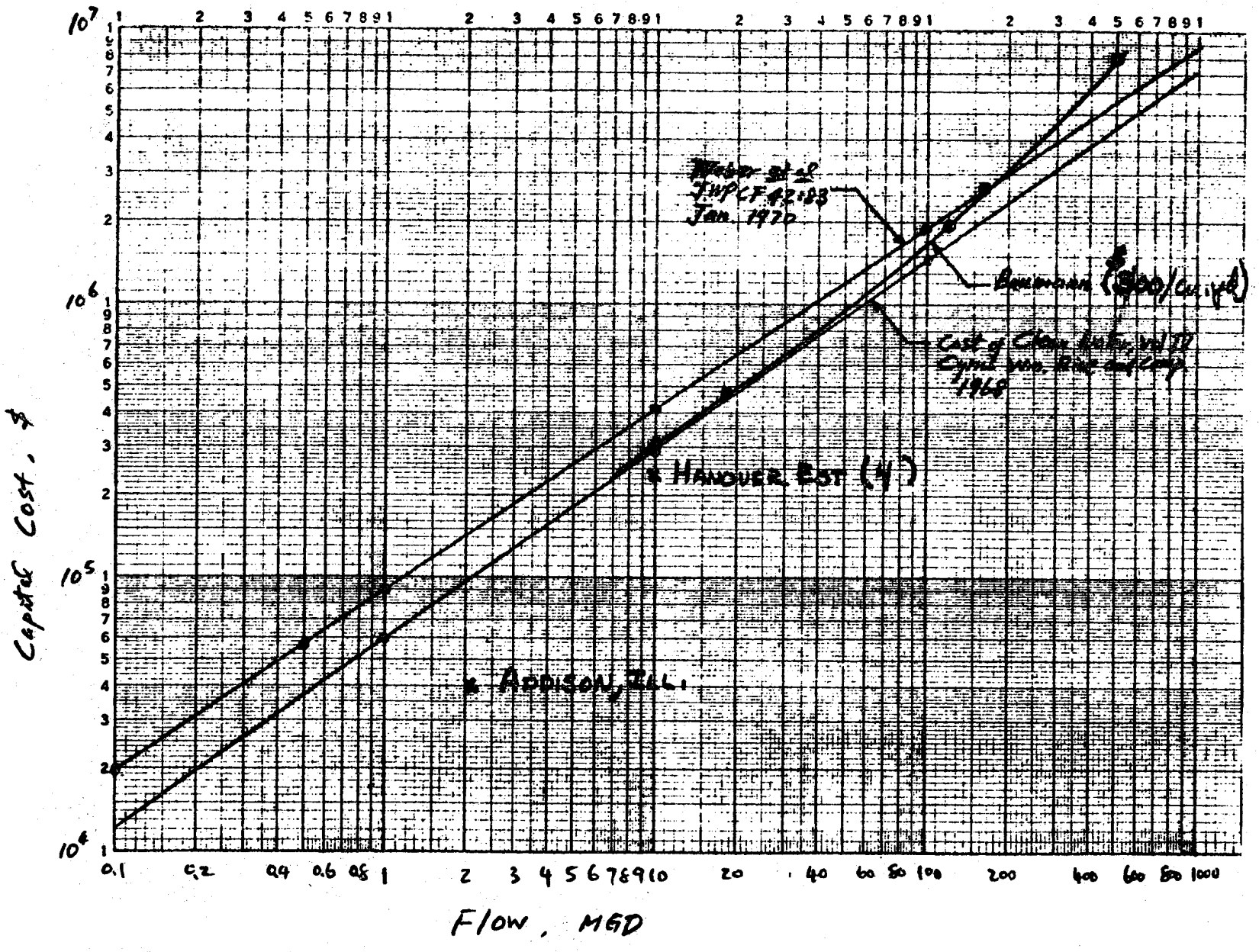


FIG 7 Capital Costs For Tertiary Treatment Filters

Table 8. Capital cost of tertiary filtration plant.

		Winnipesauksee		L-L-H		Boston	
		Mean flow 9.72MGD	Design flow 18.3MGD	Mean flow 121MGD	Design flow 164MGD	Mean flow 500MGD	Design flow 690MGD
No. of filter unit (4 filter cell per unit)		3	3	7	9	27	37
Service capacity per unit, including 4% backwash water		3.33MGD	6.35MGD	18.0MGD	18.9MGD	19.2MGD	19.4MGD
Filter area per unit at rate 4 gpm/ft^2		460 ft^2	1,100 ft^2	3,120 ft^2	3,280 ft^2	3,230 ft^2	3,360 ft^2
Equipment cost		$ 34,310	$ 50,550	102,150	107,450	$ 112,200	$ 115,650
Concrete cost	$200/cu yard	$ 46,000	$ 71,600	118,200	122,800	$ 124,000	$ 126,000
	$300/cu yard	$ 69,000	$107,500	177,500	184,000	$ 186,200	$ 190,000
Total cost/unit	$200/cu yard	$ 80,310	$122,150	220,350	230,750	$ 236,200	$ 241,650
	$300/cu yard	$103,310	$158,050	279,650	291,950	$ 298,400	$ 305,650
Total Plant cost	$200/cu yard	$240,930	$366,450	$1,542,450	$2,079,000	$6,380,000	$ 8,940,000
	$300/cu yard	$309,930	$474,150	$1,952,000	$2,630,000	$8,050,000	$11,290,000

Regretfully, operating cost data are lacking. Lynam and Bacon (4, 5) estimate the total operating cost of a 10MGD performance capacity plant based on Hanover plant data to be 2.40 cents/1000 gallons. Cost data given were:

Capital cost ----------	$240,100	(Fig. 7)
Amortized cost/MG ----------	$3.804	
Operation and Maintenance/MG -----	13.009	
Chemical/MG ----------	6.990	
Power	0.202	
Total	$ 24.00/MG	

This included prechlorination costs. Data from the author (1) would appear to indicate that an average operating cost for tertiary wastewater filtration should be in the neighborhood of 2.0-2.5 cents/1000 gal. Naylor (4) reported that the cost of the downward-flow sand filters operated for 15 years at Luton, including depreciation and running costs, including repayment of capital cost, worked out at 5.45 cents/1000 gallons. They expected upward flow filters to significantly decrease this operating cost.

9. SAMPLE CALCULATIONS AND TYPICAL PLOTS

Filter design characteristics

In the design of a filter for tertiary waste treatment, the filter characteristics which must be established include:

Filter media size and depth
(single media; graded or ungraded)
(multi-media; types, specific gravities, sizes)

Filtration rate, gpm/sq. ft.

Terminal pressure drop, ft. of water

Present design practice tends to use filter media which is too small in size and too shallow in depth, filtration rates and terminal head losses which are too low, and filtration experience which is based on potable water rather than waste filtration experience. As a result field experience available in the United States is not really sufficient to outline with the required degree of precision expected filter performance under the wide variety of waste filtration applications that are possible in tertiary treatment situations. In those situations where the actual waste to be treated exists, a series of pilot plant studies conducted over a period of several weeks can provide the information needed for optimum economical design purposes.

In designing a pilot plant study, the following points should be considered:

1. Filters should be compared on the basis of their ability to provide equivalent performance. Two or more operationally optimum filters may be said to provide equivalent performance when they produce the same quantity and quality of filtered water from the same water source during the same time period. Thus, equivalent performance

means that the quantity and quality of the filtered water produced in one day, for example, must be the same, but their media size, media depths, filtration rates and terminal head losses may be significantly different.

2. The filter which permits the equivalent performance to be achieved at least cost would represent the optimum filter design. In general, few, if any, filters have been designed which can produce equivalent performance at least cost. A recent paper by the author (1), indicates that least cost filtration of an iron suspension (Figure 8) tends to require coarse media, high filtration rates, and relatively large media depths. In Figure 8, the least cost filter design is not reached, but must be for media size greater than 1.3mm and a correspondingly higher filtration rate and media depth. Optimum design in this case provided about 30% cost savings over that obtained by standard practice design.

Ives (2) in a very recent paper, has also suggested techniques for least cost filter design. In applying these techniques using two mathematical models for iron removal and British filter cost data, Ives concluded that the least cost filter would have the following characteristics:

Media depth:	8 ft.
Media size:	1.5mm
Head Loss:	10 ft.
Filtration Rate:	7.5 gpm/ft^2

These data would tend to confirm those by Huang and Baumann (1). Ives (11) has also demonstrated that for each media grain size there is a maximum media thickness beyond which no operational optimum can be

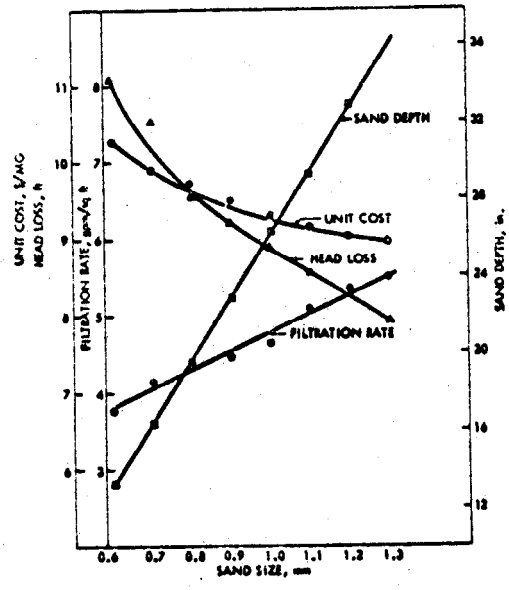


FIG. 8 Least cost design characteristics of filters for iron removal (2)

established and provision of media depth beyond this maximum merely wastes filter media and filter space and adds slightly to operational costs.

3. The length of the operational optimum filter run desired will depend on the method of filter operation -- whether manual or automatic backwash and filter operation is provided. More water will be used for backwash as the frequency of backwash increases. With manual operation, filter runs should exceed 24-30 hours. With automatic operation, shorter filter runs are economically feasible, but ordinarily filter run length should exceed 12-15 hours.

With these considerations in mind, a filter pilot plant can be designed and operated to provide the data necessary to determine the optimum operational characteristics and to approach the optimum economical characteristics of a tertiary waste treatment filter. Such a filter pilot plant might involve the operation of three or four batteries of 4 filters per battery (diameter at least 50 x largest filter media size to eliminate scale effect) in a series of runs. The filters in each battery would be filled with 2", 8", 16", and 32" of filter media. The run characteristics might be designed in accordance with the following phases:

Phase A Study of Single-Media Size Effects on Performance.

Filter Batteries A, B, and C contain media sizes 1, 2, and 3 respectively.

Filter runs 1, 2, 3, etc. using all batteries at flow rates of 2, 4, 6, 8 gpm/ft^2 respectively, observing effluent quality and head loss in each filter.

Phase B Using Media Size from Phase A which promises desired operational characteristics, study effect of flow rate on performance.

Filter Batteries A, B, and C all contain same media size indicated in Phase A.

Filter runs 1, 2, and 3 made using batteries A, B, and C operated at filtration rates of 2, 4, 6 or 4, 6, and 8 gpm/ft^2 respectively.

Phase C Study of single media filters vs. dual (or multi-) media filters using media size and flow rates which promises desired operational characteristics observed in phases A and B.

Filter Batteries A, B, and C contain different media.

Battery A - Same media as Phase B.

Battery B - Top media, coal 2 x media size in Battery A, bottom media same as Battery A.

Battery C - Top media, coal 1-1/2 x media size in Battery A, bottom media same as Battery A.

Figures 9 through 17 show some of the results of a series of pilot plant runs made with such a battery of filters in the removal of suspended solids from a typical trickling filter plant final effluent at Ames, Iowa. Figure 9 shows the filter media sizes and depths in the filter batteries in the phase A runs. Three media sizes were used (passed/retained, 18/20 or 0.92mm, 25/30 or 0.65mm, 30/35 or 0.55mm). Figure 10 shows the suspended solids in the influent and effluent and head losses observed in a run at 6 gpm/sq. ft. The following conclusions can be drawn from these results:

1. The three filters gave approximately the same effluent quality, with the 0.55mm sand providing the best quality, 0.65 sand the next, and 0.92mm the next best.

2. The head losses in the 0.55 and 0.65mm sand increased so fast as to preclude the use of these media sizes in this application. The head loss in the 0.92 mm increased rapidly, but not nearly so far as for the other media.

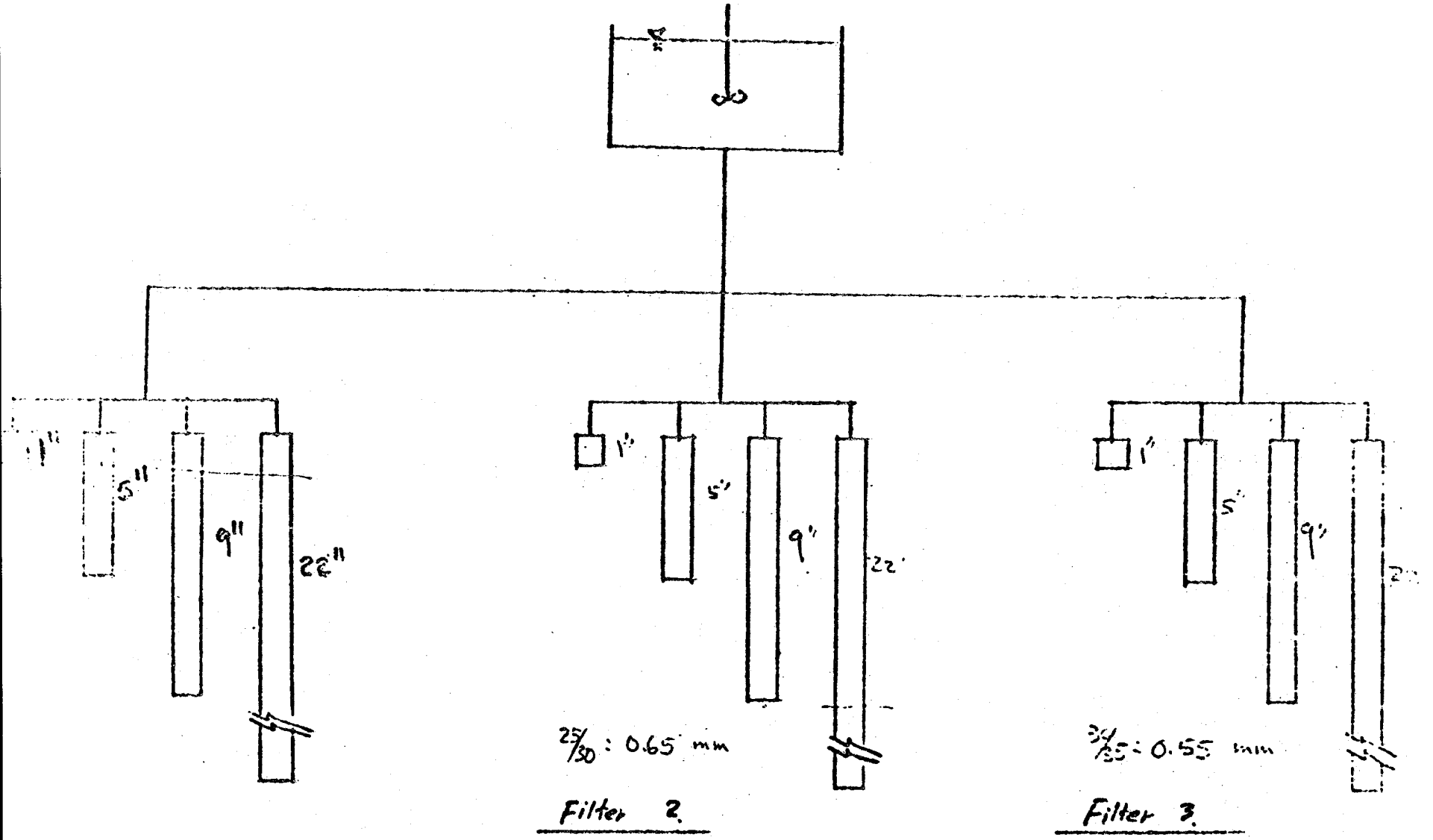
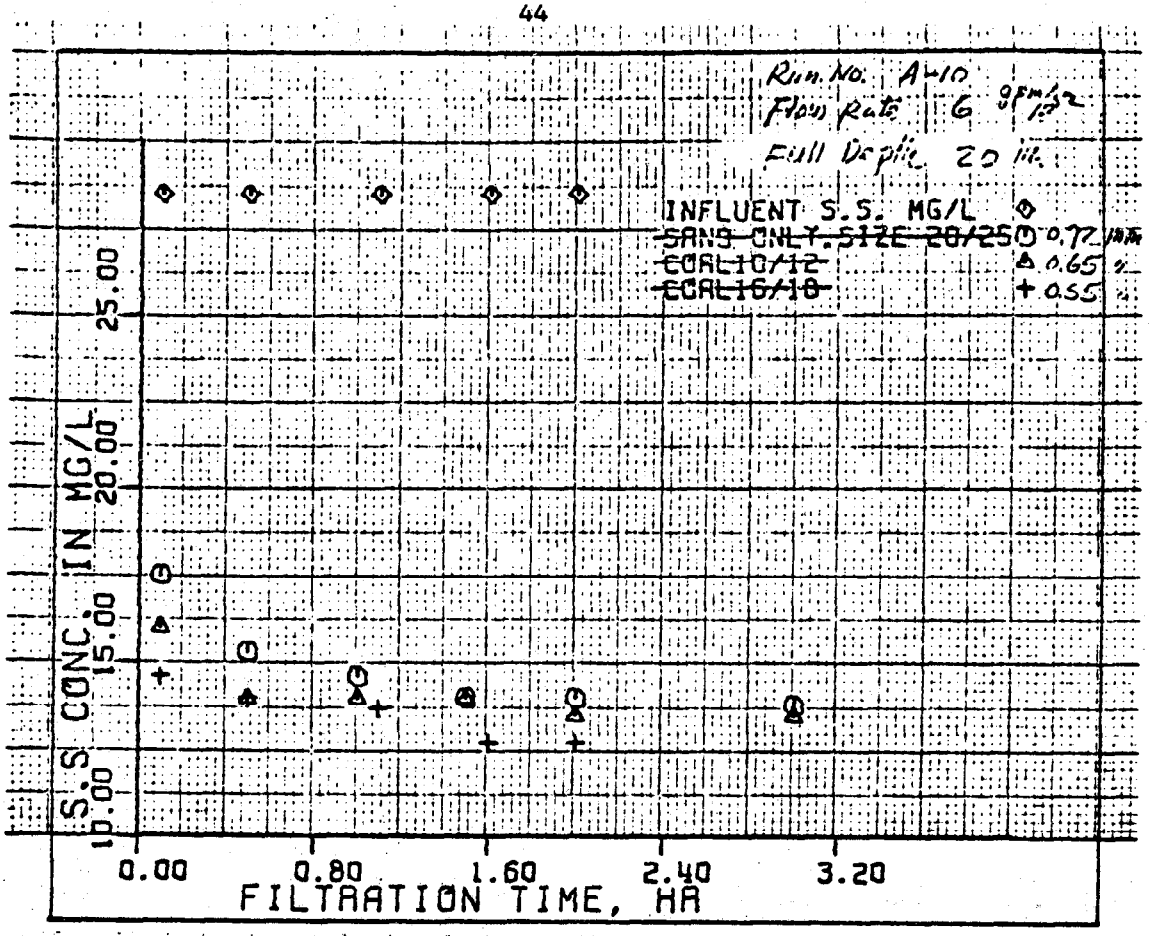


FIG 9. FILTER CHARACTERISTICS, PHASE A.



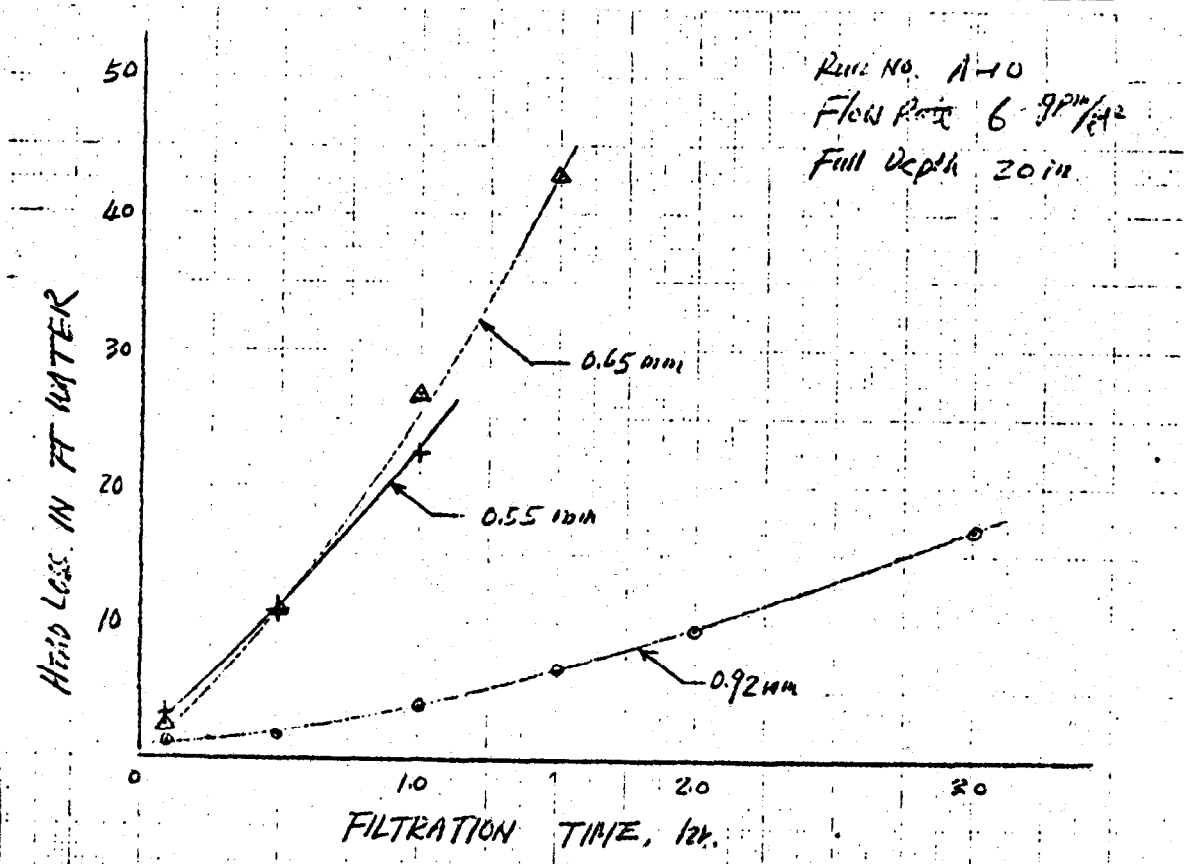


FIG. 10 Phase A, Q = 6 gpm/sq ft, 20-in. deep media.

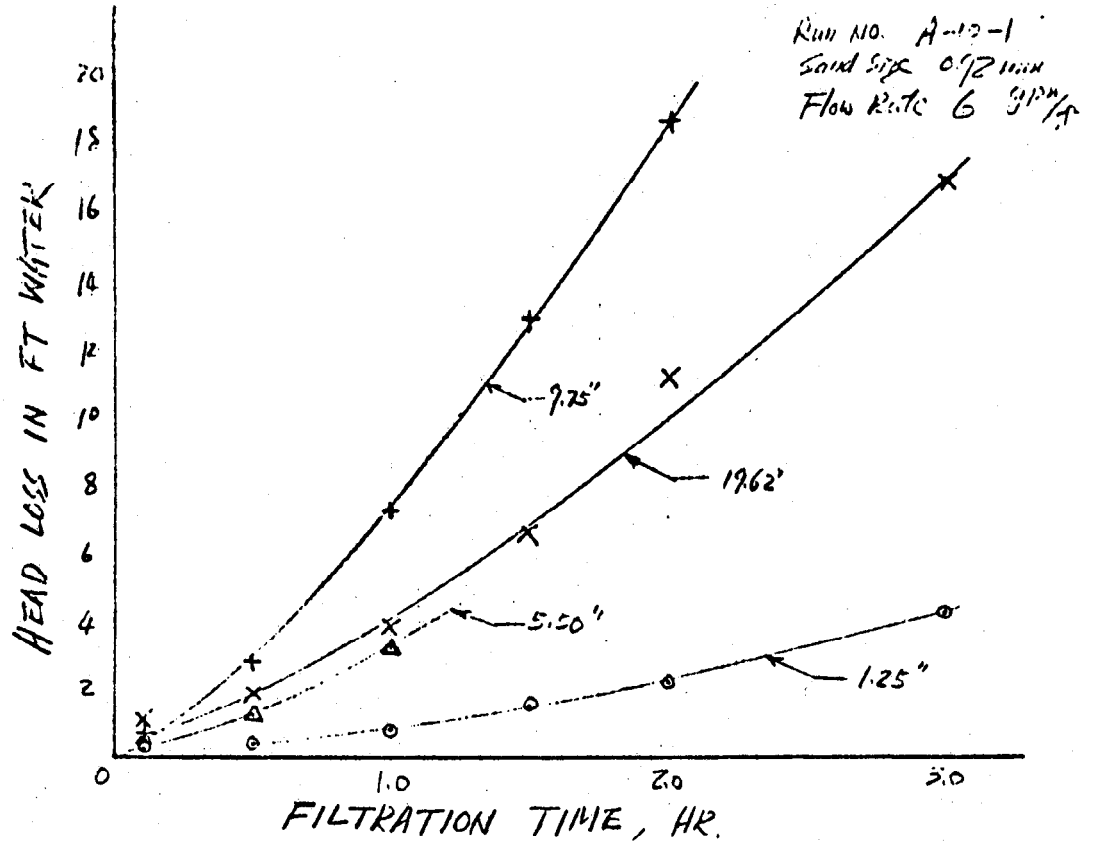
Figure 11 shows the results using the 0.92mm sand battery of 4 filters operated at 6 gpm/sq. ft. The following conclusions can be drawn from these results:

1. The effluent quality improves significantly with media depth using this sand size. Thus, increasing the media depth should provide quality equivalent to or better than that obtained with the smaller sized media and with little increase in head loss.
2. Further tests should be made with 0.92mm or larger media.

Figure 12 shows the various filter operating characteristics in one Phase B run made at filtration rates of 2, 4, and 6 gpm/sq. ft. Figure 13 shows the results of one run when the influent suspended solids were at 34 mg/1. The results lead to the following conclusions:

1. The effluent quality is best with the lowest filtration rate (5 mg/1) and becomes poorer with increased filtration rate.
2. The head loss is least, as might be expected, for the 2 gpm/sq. ft. filtration rate, but the 4 and 6 gpm/sq. ft. head loss curves are approximately the same.
3. With this filter media, a filtration rate slightly less than 2 gpm/sq. ft. and a media depth of 30-40 inches could be expected to provide excellent suspended solids removal and a 6-hour filter run with a 10 ft. terminal head loss.

Figure 14 shows the results of a second phase B run in which the influent SS were only 22 mg/1. In this run, all three filtration rates provided an effluent SS level of about 4 mg/1, again with the lowest rate providing the best quality.



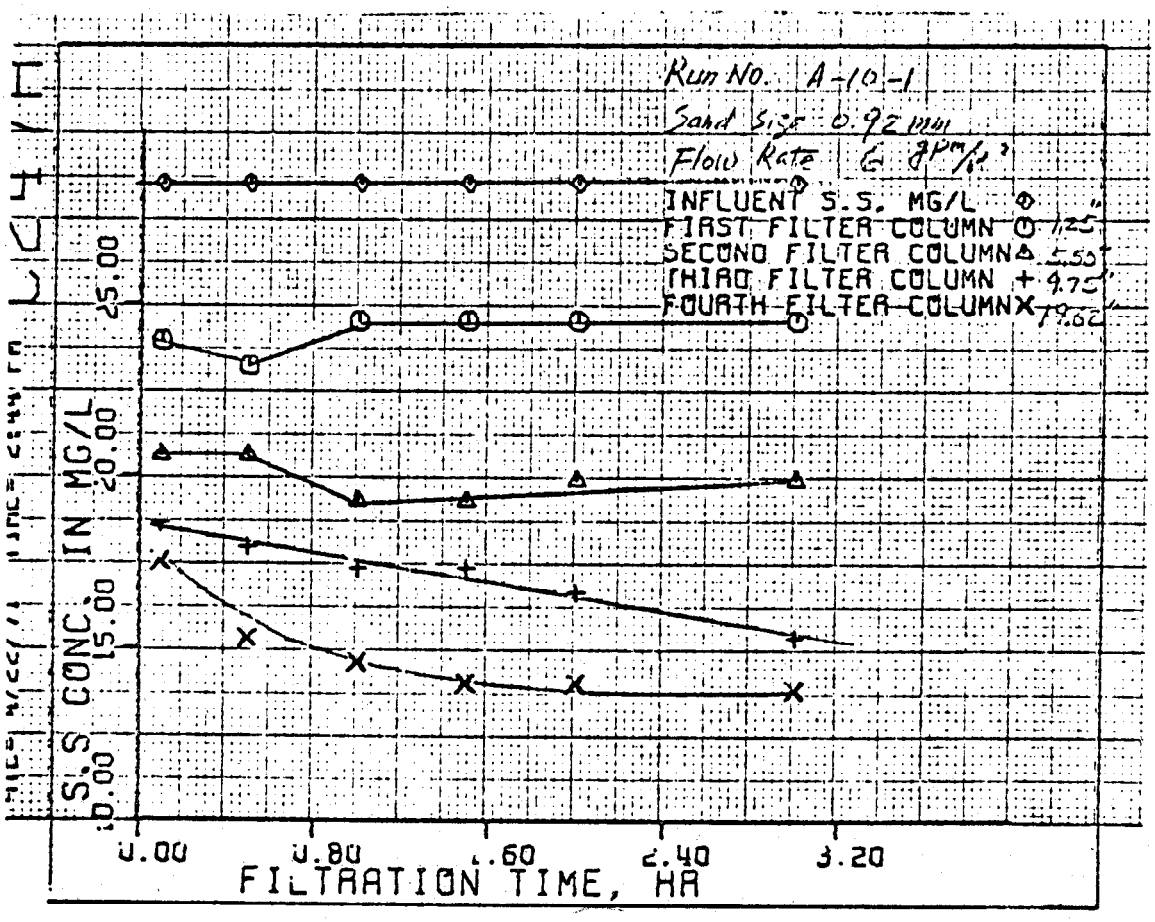
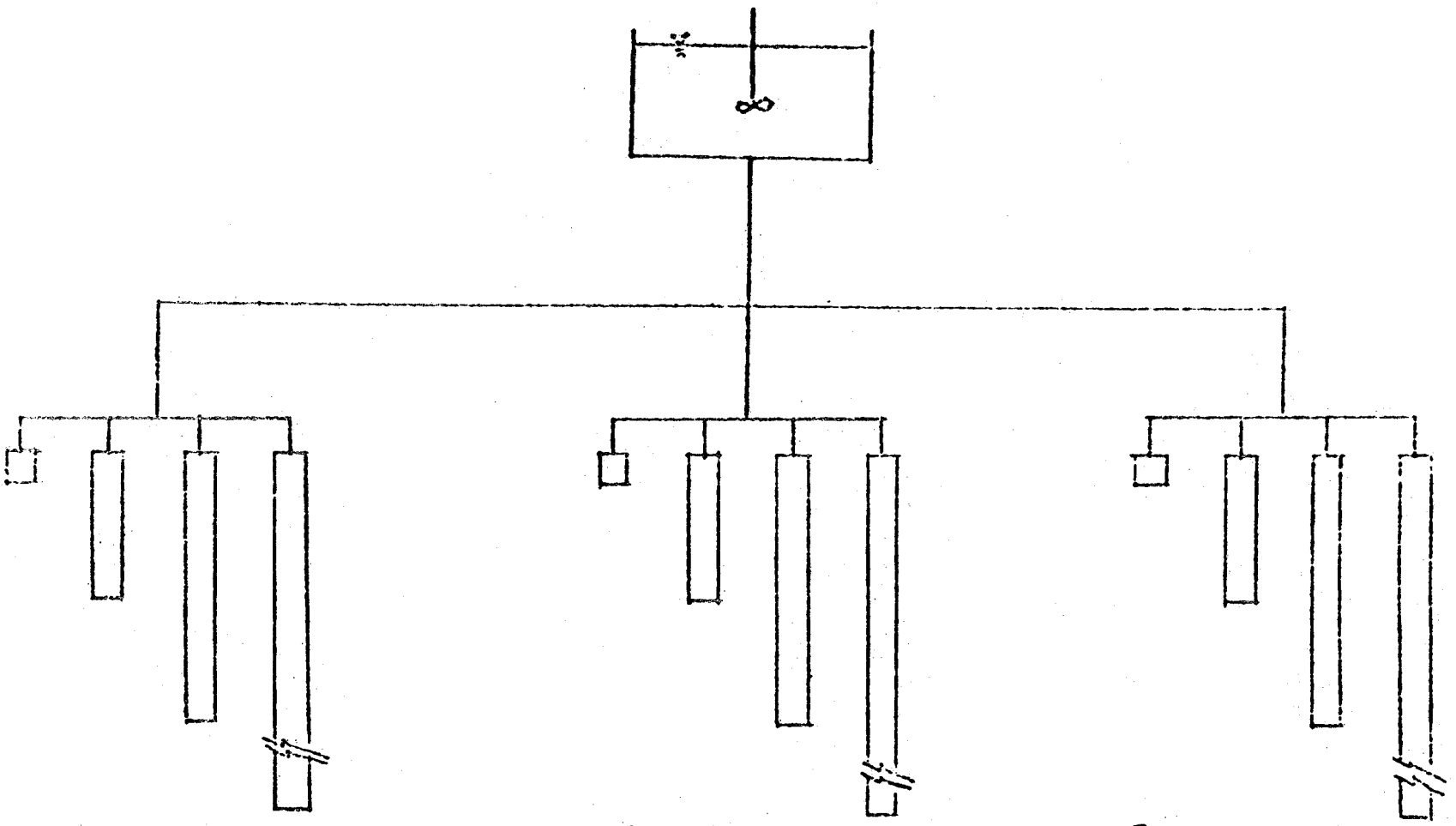


Fig 13 Phase A, 0.92 mm sand, Q = 6 gpm/sq ft., Variable media depth



Filter 1

Sand Size 0.92 mm, 1/20

Flow Rate 2 gpm/ft²

Filter 2

Sand Size 0.92 mm, 1/20

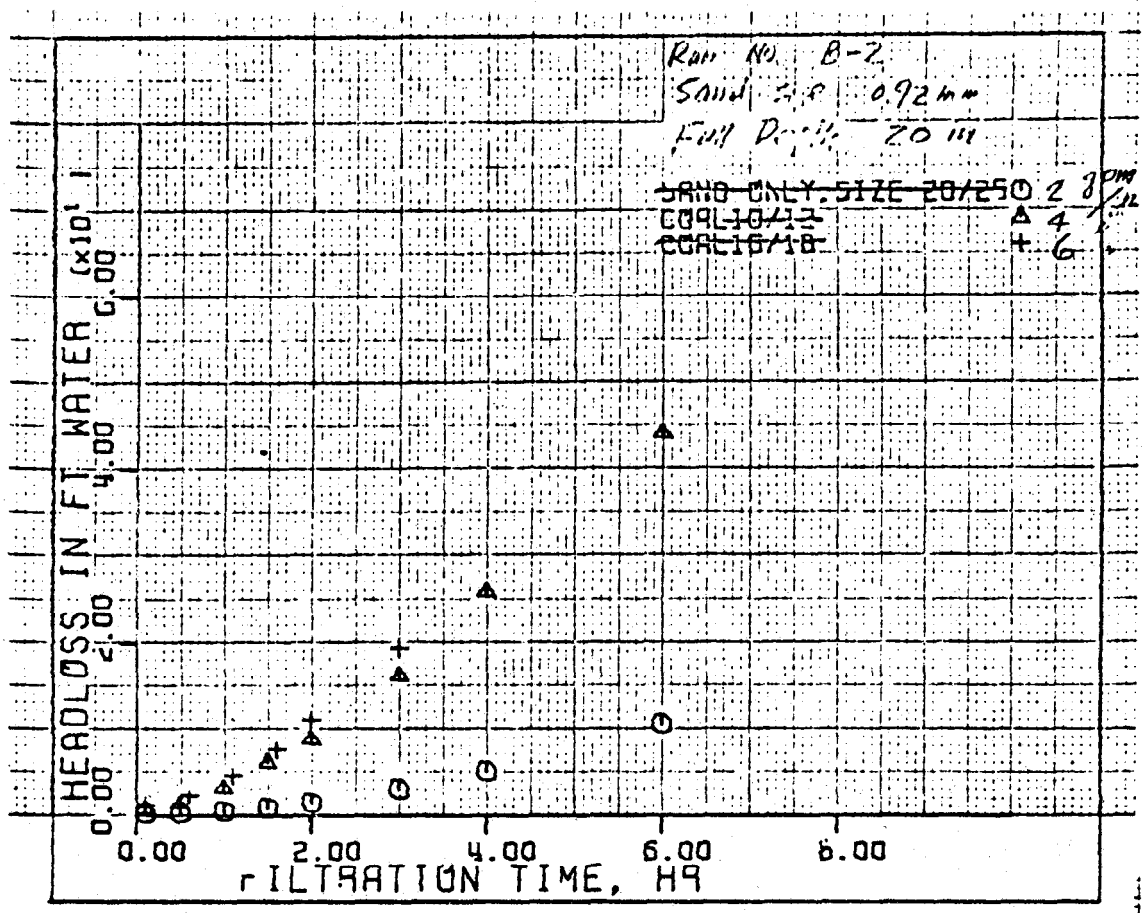
Flow Rate 4 gpm/ft²

Filter 3

Sand Size 0.92 mm, 1/20

Flow Rate 6 gpm/ft²

Fig 12 FILTER CHARACTERISTICS, Phase B



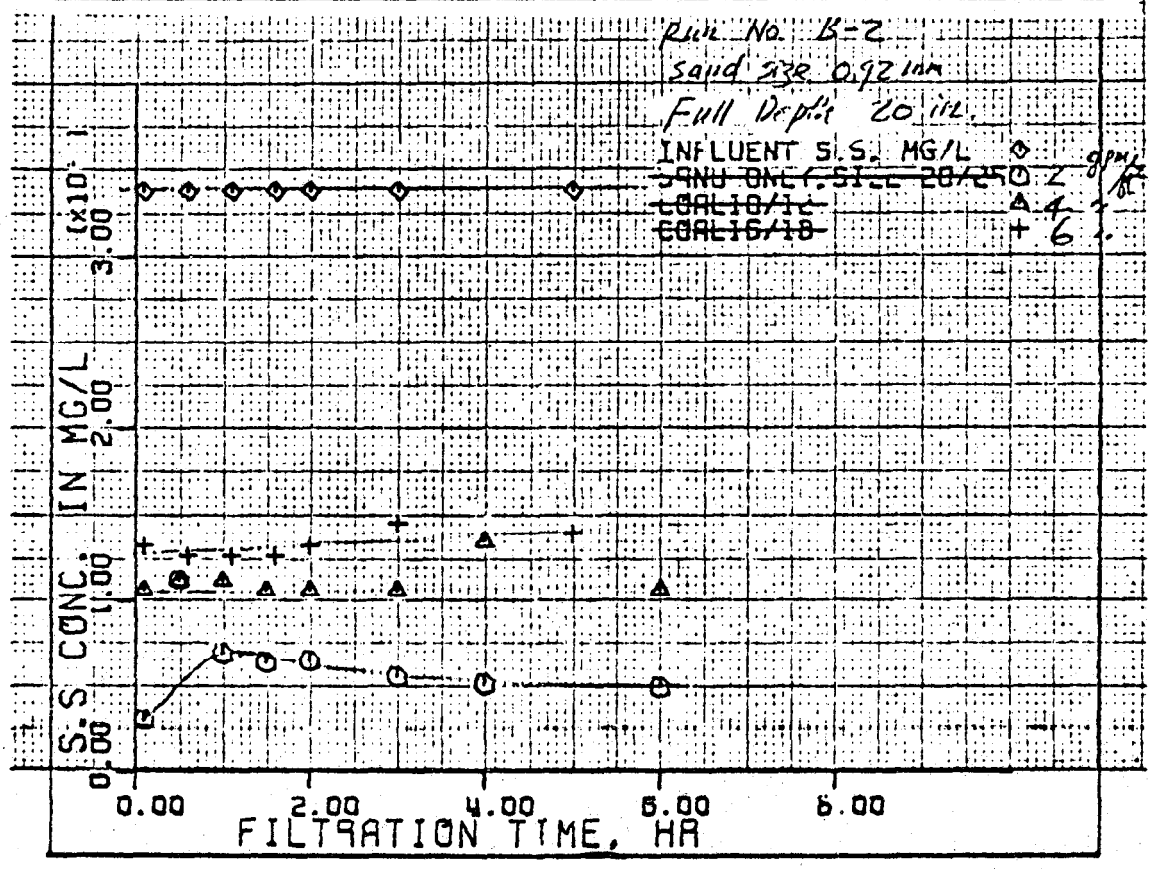


FIG 13 Phase B, Q = variable, 20-in. deep media.

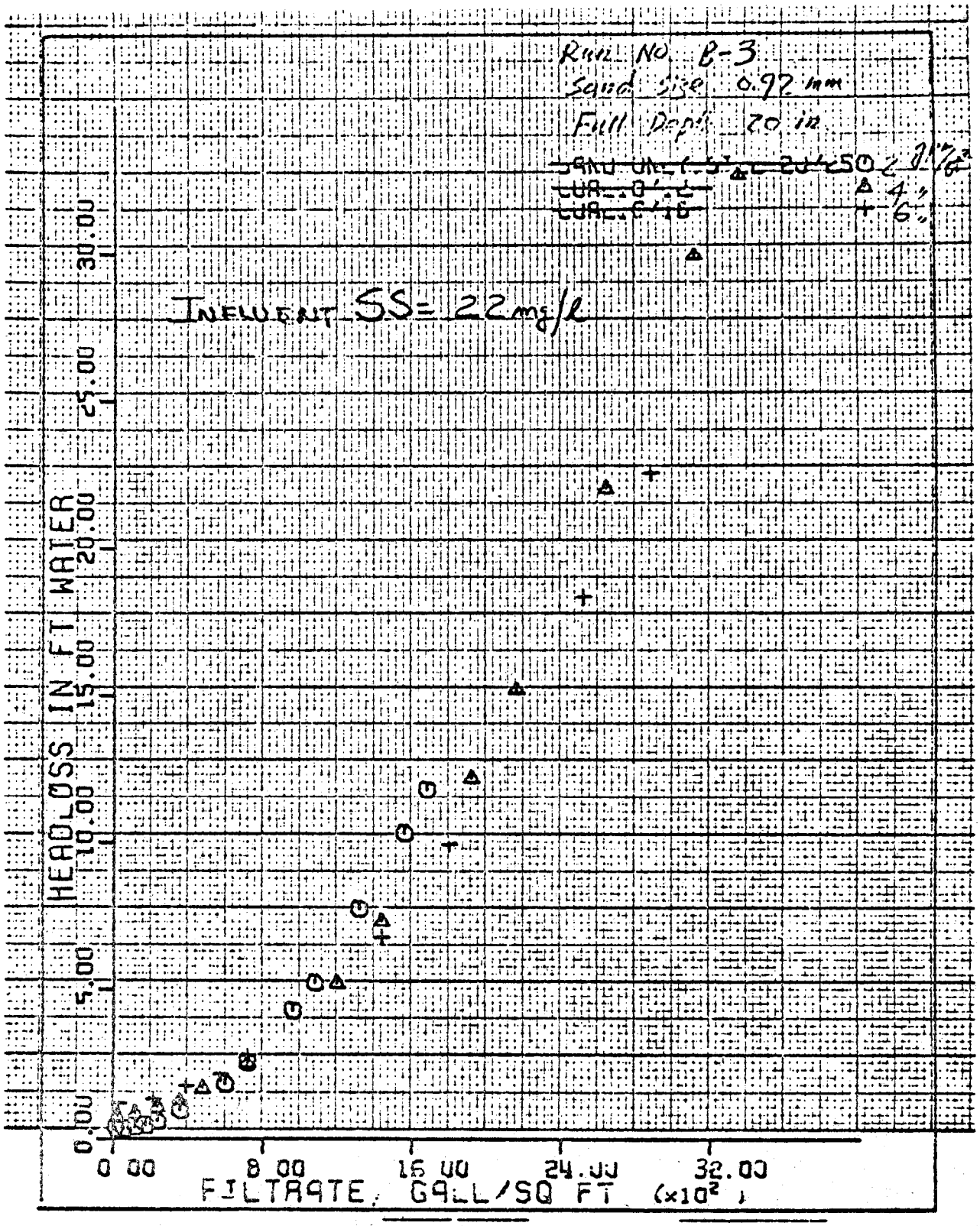


Fig 14 Phase B, 0.92mm sand media, Influent SS= 22mg/l, Q= variable.

These results indicate that all three filtration rates provide about the same head loss per gallon of filtrate. Here, the 2 gpm filtration rate would provide a 13-hour run to a terminal head loss of 10 ft. of water.

Phase C runs were then made with the filter characteristics shown in Figure 15. Media sizes were:

Filter Battery 1,		16/18 sand (1.09mm)
Filter Battery 2,	Top:	10 inches of 10/12 coal (1.84mm)
	Bottom:	10 inches of 20/25 sand (0.75mm)
Filter Battery 3,	Top:	10 inches of 16/18 coal (1.09mm)
	Bottom:	10 inches of 20/25 sand (0.75mm)

Figure 16 shows the results of a run of 2 gpm/sq. ft. when the influent SS were 34 mg/l (the same as in Fig. 13). The results lead to the following conclusions:

1. All three filters provided essentially the same quality of effluent (SS - 3 mg/l), with the 1.09 coal-0.75mm sand tending to very slightly better quality.
2. The 1.84mm coal-0.75mm sand filter provided the lowest head loss, such that a 24-hour run could easily be obtained with a head loss of well under 10 ft.

Figure 17 shows the results from the 4 filters in the 1.84mm coal-0.75mm sand filter battery. Here, most of the SS removal is taking place in the top 7 inches of the coal, with little SS removal and head loss occurring in the bottom 13 inches of media. Increasing media depth would provide some increased removal but little additional head loss. Here, use of a small (6") depth of 0.55mm garnet sand would provide further improved quality at little increase in head loss.

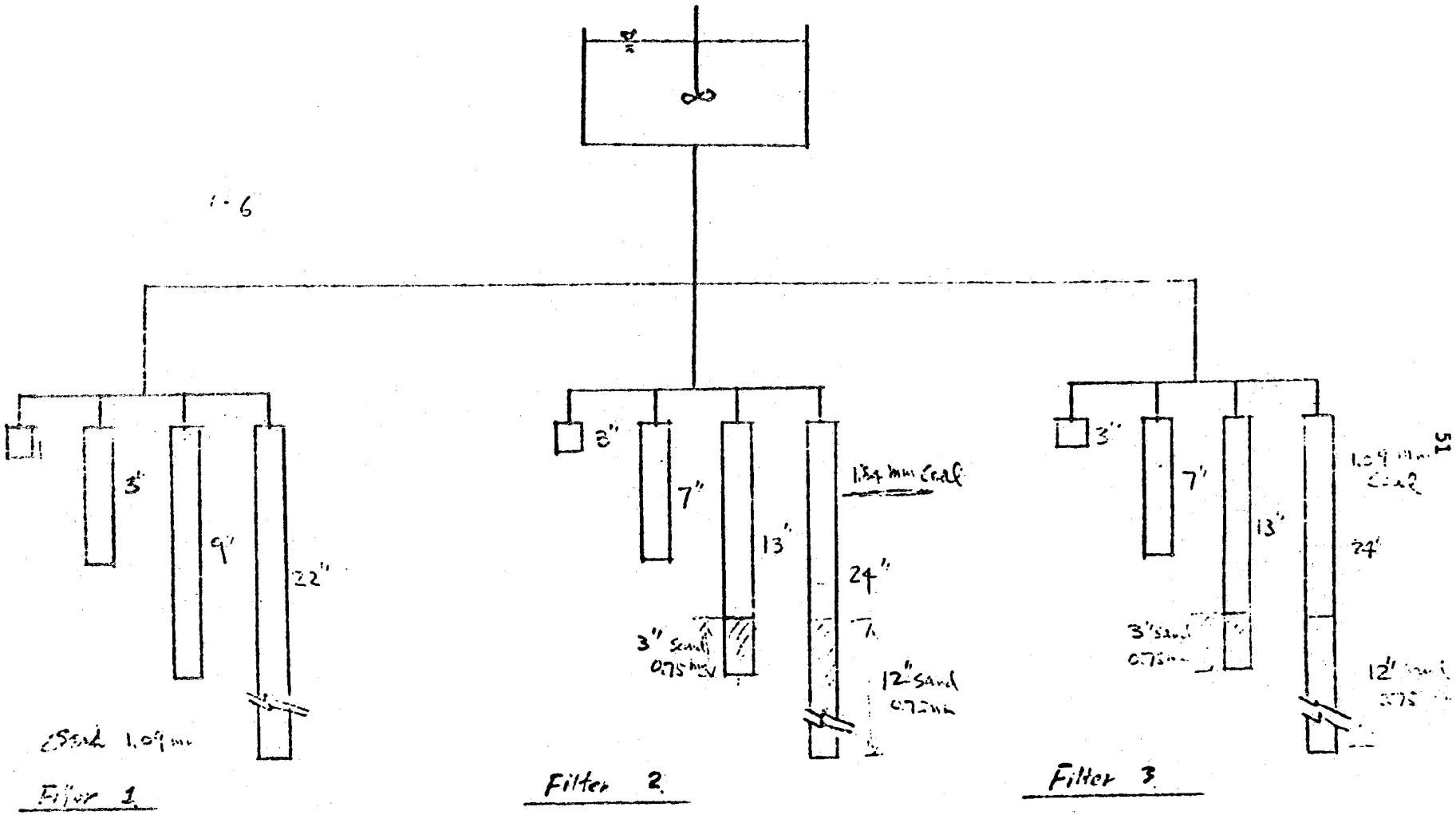
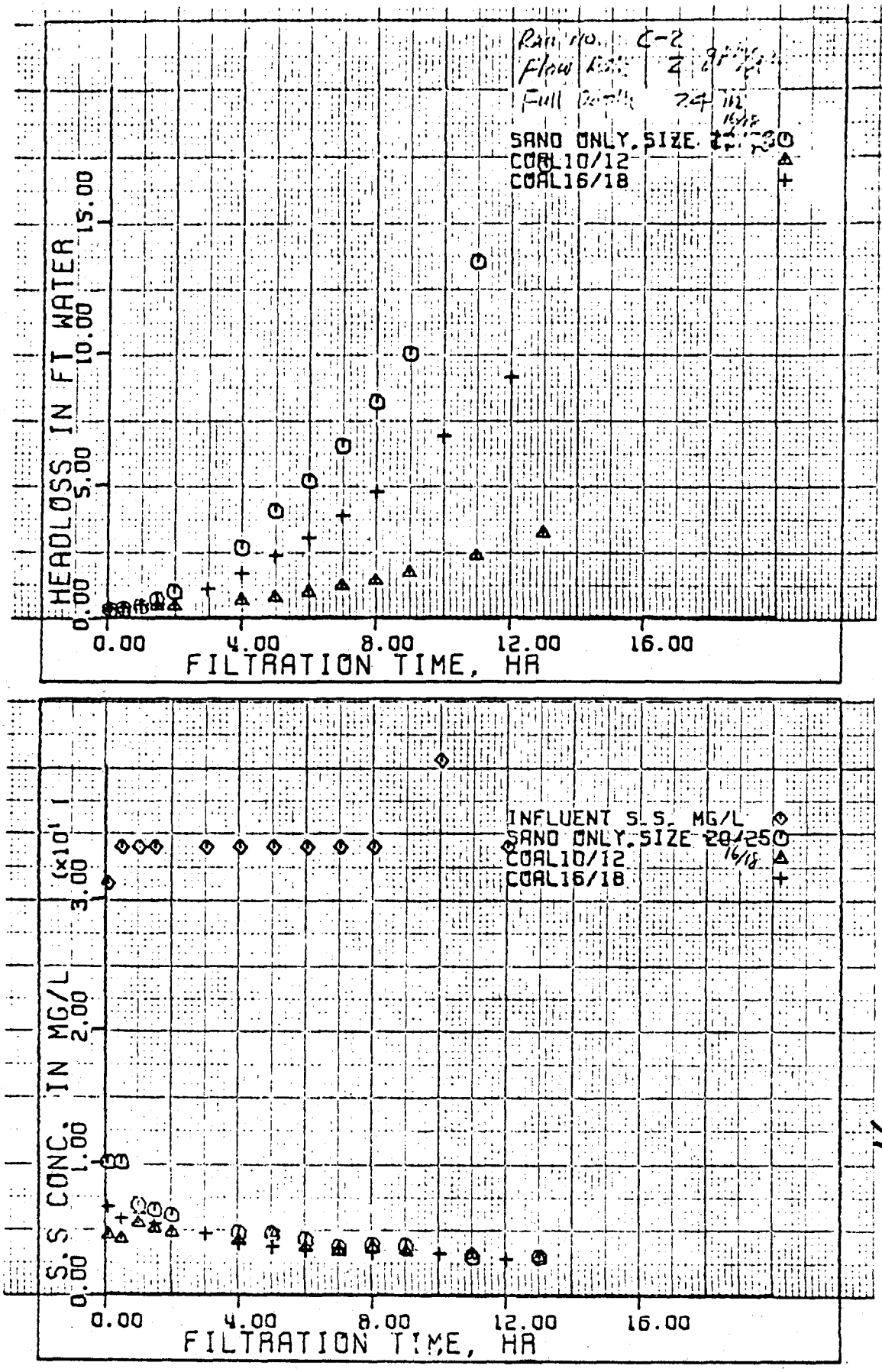
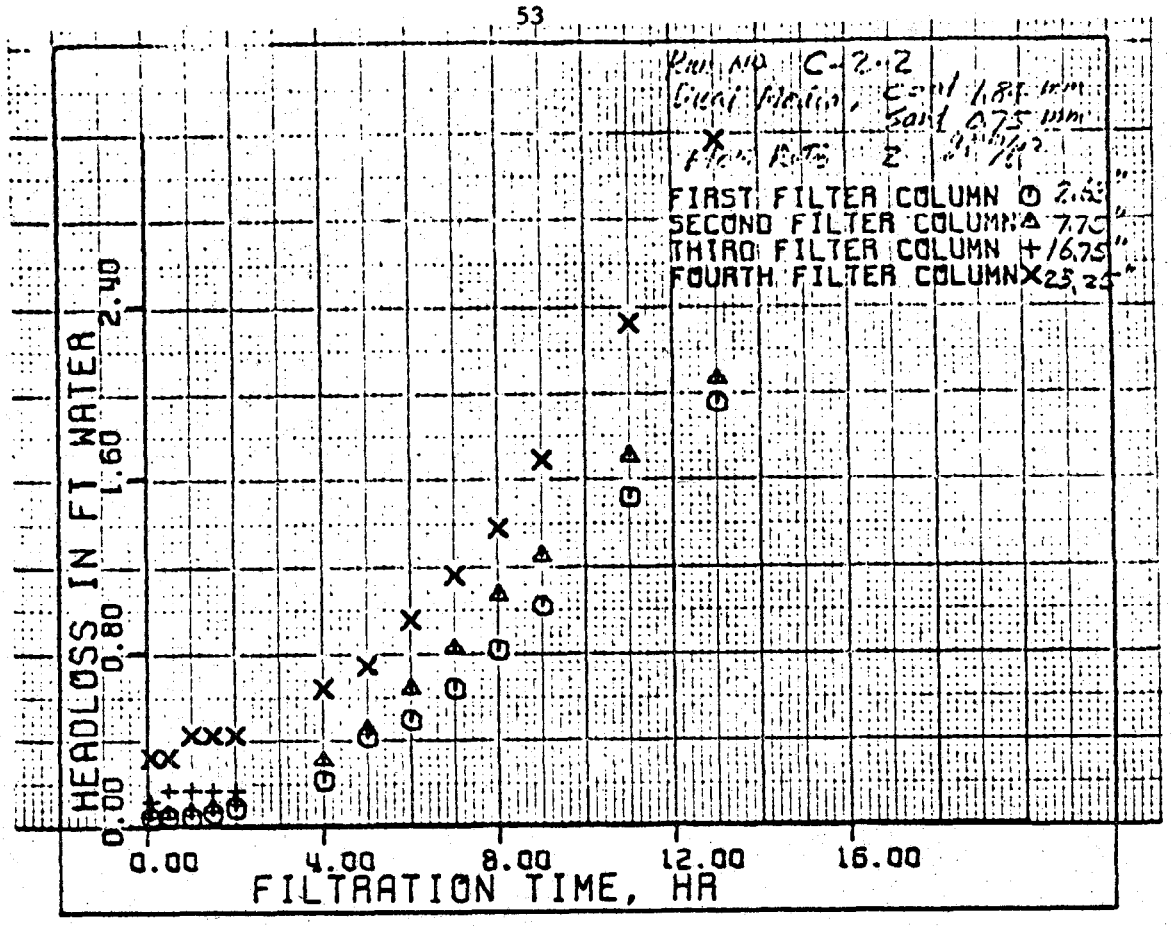


FIG 15. FILTER CHARACTERISTICS, PHASE C.



16
FIG 8. Phase C, Q = 2gpm/sq ft, 20-in deep media



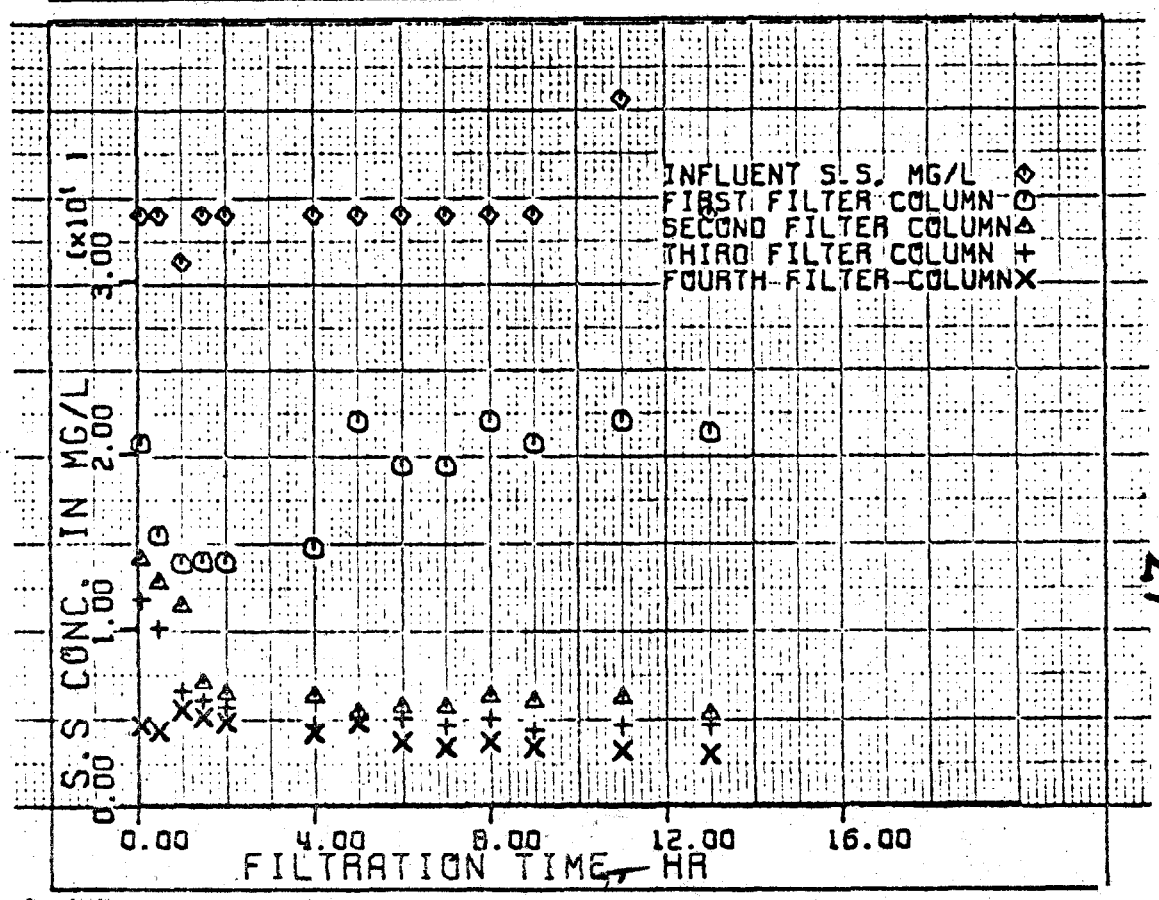


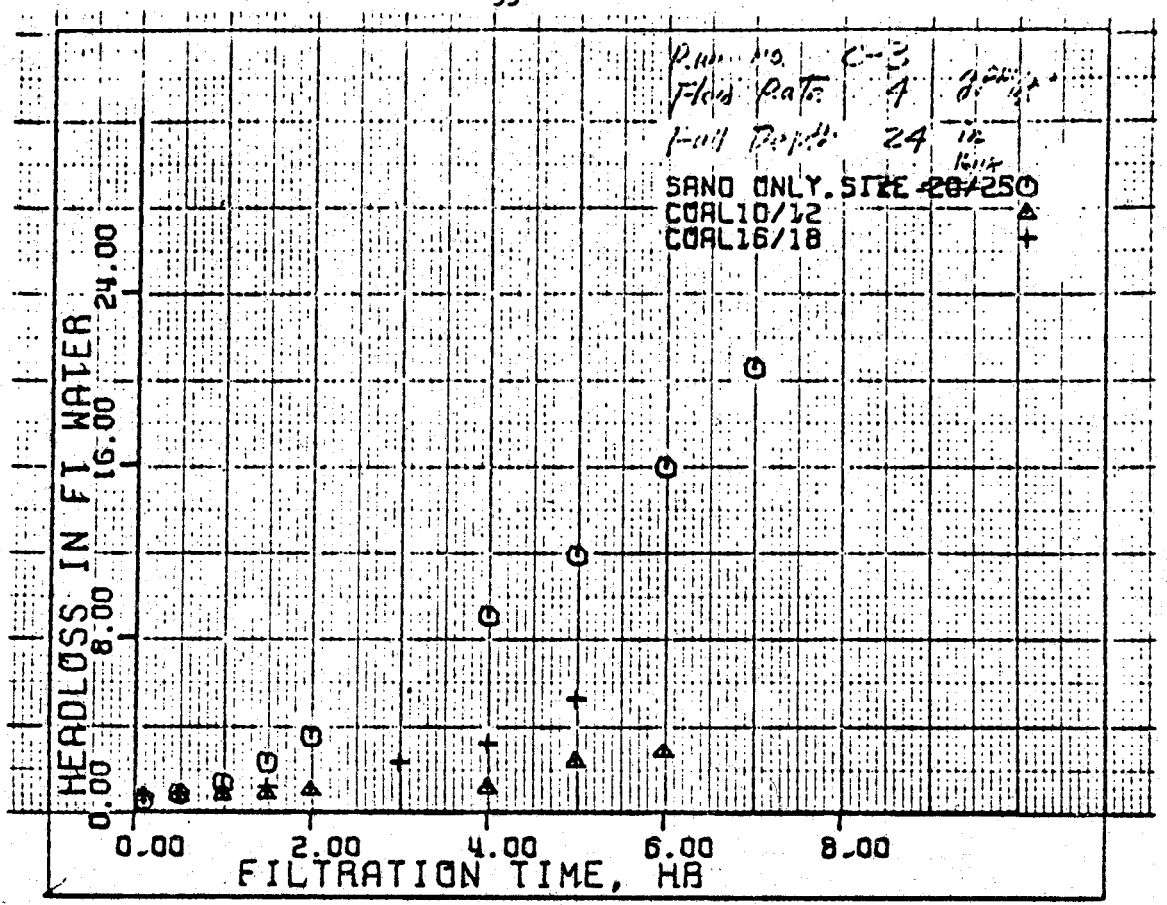
FIG. 17 Phase C, 1.84 mm coal – 0.75 mm sand media, Q = 2 gpm/cg ft, depth varied.

Figure 18 shows the results of a phase C run at 4 gpm/sq. ft. when the influent SS were at 45-47 mg/l. Here again, all three filters provided about the same effluent quality (8-10 mg/l), and the 1.84mm coal-0.75mm sand filter provided significantly reduced head loss. This filter should be able to provide a 24-hour run, even at the 4 gpm/sq. ft. filtration rate. Figure 19 shows the results from the four 1.84mm coal-0.75mm sand filters in this run. The results indicate again that most of the solids removal was in the top coal layer. Once again, use of a 6" layer of a finer (0.55mm) garnet sand should provide improved filtrate quality with little additional head loss.

On the basis of these results, a Phase D series of runs might be made to evaluate effect of filtration rate (2, 4, 6 or 4, 6, 8 gpm/sq. ft.) on filtration results using a 3-media filter. The media might be a compatible series of:

Top:	10-14 inches of 1.84mm coal
Middle:	10 inches of 0.75mm silica sand
Bottom:	6 inches of 0.40-0.50mm garnet sand

The results of such runs should provide all data needed for fixing a "functional" final filter design.



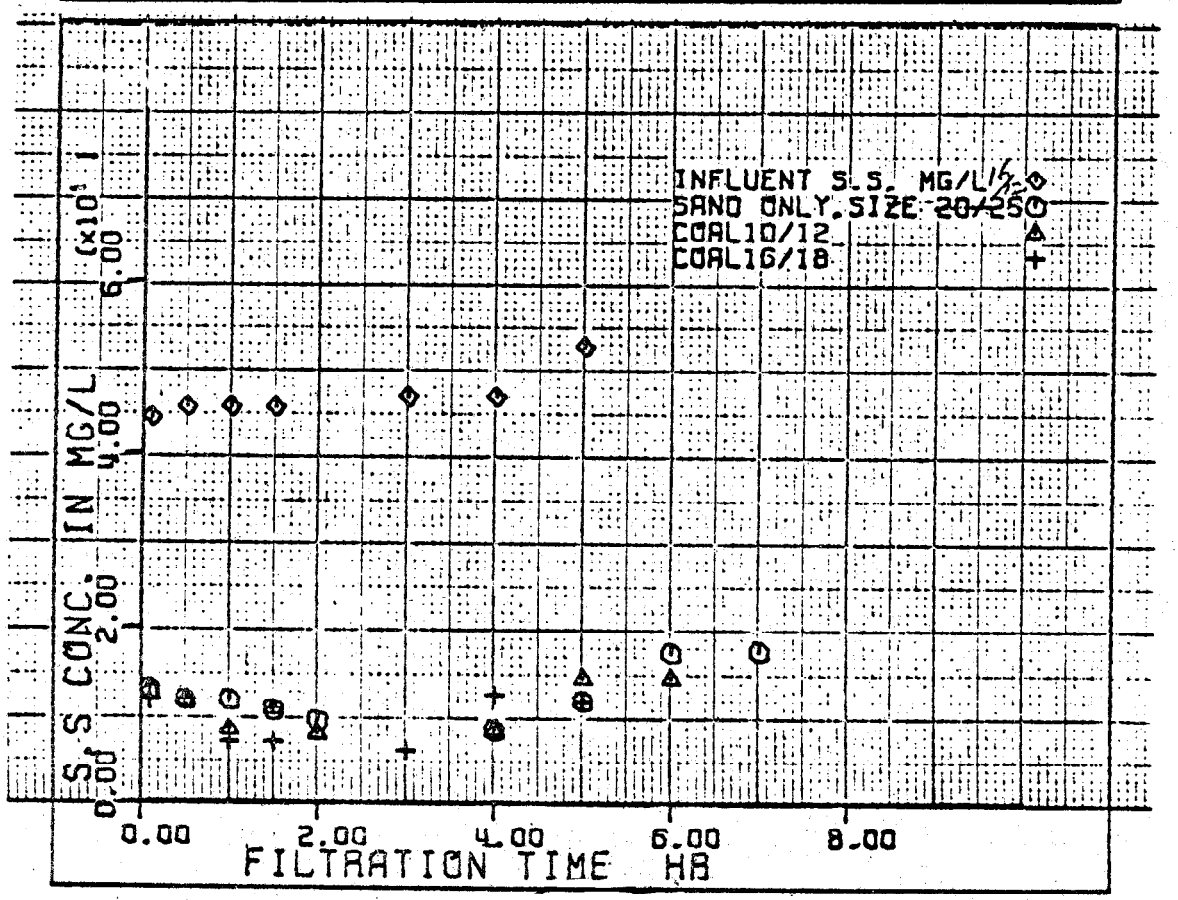
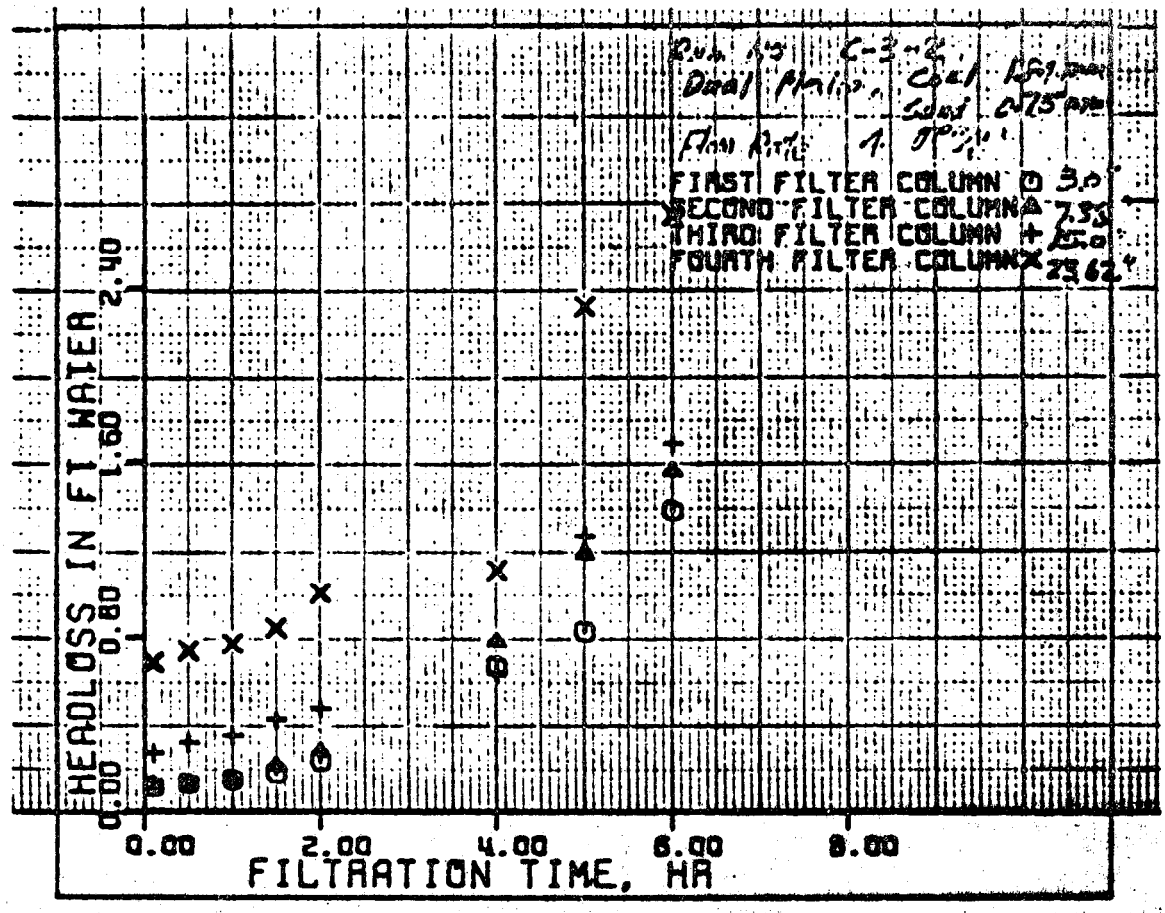


Fig. 18. Phase C, Q = 4 gpm, log μT, 20-in. dual media



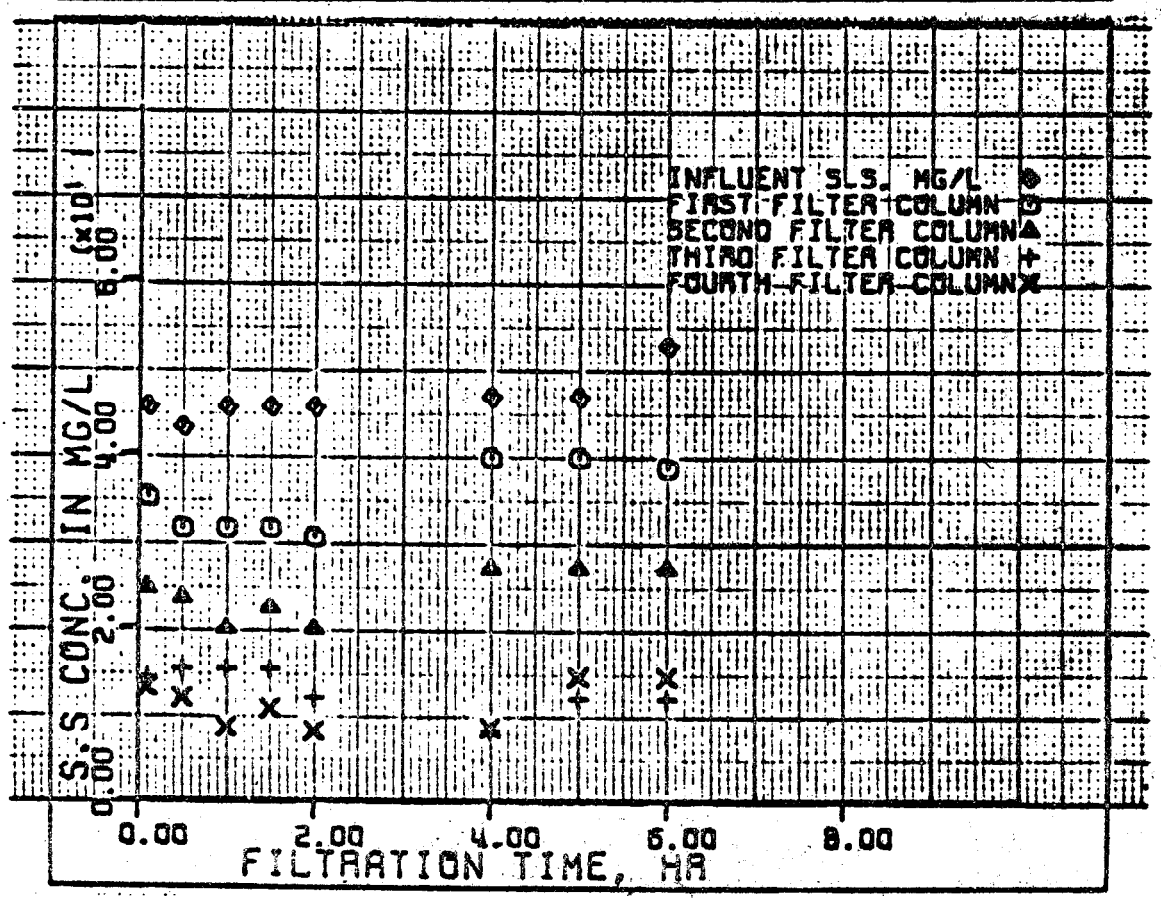


FIG A Phase C, 1.84mm coal · 0.75mm sand filter, depth variable, Q = 44 pm/sq ft.

10. SCALE UP FACTORS

Experience to date with filtration indicates that pilot-plant filtration results can be extrapolated directly to any size plant desired. Pilot plants themselves, however, require that the diameter of a cylindrical pilot filter be at least 50 times the diameter of the largest media size. Thus, the minimum pilot filter should have a diameter of at least (3mm anthrafilt x 50 = 150mm or 150/25.4 - 5.9 inches) 6 inches. The effects of media size and depth, flow rate, head losses, and filtering efficien of a 6-inch diameter filter will be the same as on an 850 sq. ft. filter, the recommended maximum size filter. Thus, scale up is not a problem in filtration.

11. PROCESS DEPENDABILITY

The process of filtration is one which is unusual in that it will never, if properly designed, provide a grossly inferior quality of effluent. In general, if a filter can provide the desired effluent quality under normal conditions, upsets in pretreatment processes will provide shorter filter runs and not significantly poorer effluent quality. Thus, if under normal conditions the effluent SS are running at 18 mg/l and suddenly increase to a level of 30 or 40 mg/l, the principal effect will be a significant decrease in run length but a relatively lesser increase in effluent SS.

12. MATRIX SUMMARY*

	Target Year		
	1975	1990	2020
Scale-up	No problem	No problem	No problem
Dependability Over Time	1	1	1
Flexibility	1	1	1
Tolerance to Influent Variability	1	1	1
Effluent Variability	1	1	1
Ancillary Processes Required	None	None	None
Costs	2	1	1
Process Efficiency	1	1	1
Operational Difficulty	Very low	Very low	Very low

* 1 best - 5 worst

Equipment Cost

Equipment Items	Winnic		LLH		Boston	
	Menu 460 ft²	Design 1,100 ft²	Menu 3,000 ft²	Design 3,160 ft²	Menu 3,240 ft²	Design 3,240 ft²
1. Cell isolation valves	2,500	4,750	10,250	12,000	12,000	12,000
2. Rotary surface washers	3,900	6,250	12,250	12,500	13,000	13,000
3. Cross walk	760	850	1,100	1,150	1,200	1,250
4. Wash troughs	2,900	4,500	16,000	16,500	17,000	17,500
5. Underdrain system parts	2,800	5,200	16,400	17,200	17,600	18,000
6. Filter media	700	1,500	5,600	6,000	6,500	7,000
7. Air wash-blower and valves	3,800	5,700	15,000	15,600	16,200	16,800
8. Inlet distribution	2,400	3,100	5,700	6,500	7,300	8,000
9. Central column and valves	13,800	17,800	18,600	19,200	20,000	20,500
10. Cross walk	750	900	1,100	1,200	1,300	1,400
11 backwash pump	—	—	300 x 1/2 = 150	300 x 1/3 = 100	300 x 1/3 = 100	300 x 1/3 = 100
Total Equipment cost/unit	34,310	50,550	102,150	107,950	112,200	115,650

Appendix D.

Concrete Cost

	Willie		LLH		Boston	
	Mean 460 ft²	Design 1,100 ft²	Mean 3,120 ft²	Design 3,280 ft²	Mean 3,330 ft²	Design 3,360 ft²
Dimension	21.4'x21.4'	30'x36.6'	50'x62.5	50'x65.6'	50'x66.6'	50'x67.2
Filter floor (ft³)	(21.4+14)x (21.4+5)x 1.5 = 1,400	(30+14)x (36.6+5)x 1.5 = 2,740	(50+14)x (62.5+5)x 1.5 = 6,470	(50+14)x (65.6+5)x 1.5 = 6,800	(50+14)x (66.6+5)x 1.5 = 6,900	(50+14)x (67.2+5)x 1.5 = 7,000
Cell walls	2x21.4x1x 11 = 472	2x36.6x1x 11 = 806	2x62.5x1x 11 = 1,375	2x65.6x1x 11 = 1,440	2x66.6x1 x11 = 1470	2x67.2x1 x11 = 1,480
Filter walls	3x(21.4+1) x1x24 = 1,610 3x(21.4+15) x1x24 = 2,620	3x(36.6+1)x 1x24 = 2,700 3x(30+15) x1x24 = 3,240	3x(62.5+1) x1x20 = 3,800 3x(50+15) x1x20 = 3,900	3x(65.6+1) x1x20 = 4,000 3x(50+15) x1x20 = 3,900	3x(66.6+1) x1x20 = 4,010 3x(50+15) x1x20 = 3,900	3x(67.2+1) x1x20 = 4,060 3x(50+15) x1x20 = 3,900
Effluent Box	3.5x6.5x1.5 = 28.8 6.5x13x1 = 84.5	4x10x1.5 = 60 10x13x1 = 130	3x5x9x1.5 = 206 3x9x9x1 = 243	3x5x9x1.5 = 206 3x9x9x1 = 243	3x5x9.5x1.5 = 216 3x9.5x9x1 = 257	3x5x10x1.5 = 225 3x10x9x1 = 270
Total Concrete (ft³)	6,216	9,676	15,994	16,589	16,753	16,935
Cost, $200/yard³	$ 46,600	$ 71,600	$ 118,200	$ 122,800	$ 124,000	$ 126,500
Cost, $500/yard³	$ 69,000	$ 107,500	$ 177,500	$ 184,000	$ 186,200	$ 190,000

REFERENCES

1. Ives, K. J. and A. Gur. Research on optimization of filtration. University College London. (Personal communication through paper preprint.) May, 1971.

2. Huang, Jerry Y. C. and E. Robert Baumann. Least cost sand filter design for iron removal. Journal San. Engr. Division, ASCE, 97, No. SA2, 171-190, April, 1971.

3. Hsuing, K. and J. L. Cleasby. Prediction of filter performance. Journal, San. Engr. Division, ASCE, 94:SA6, 1043, December, 1968.

4. Naylor, A. E., S. C. Evans, and K. M. Dunscombe. Recent developments on the rapid sand filters at Luton. Water Pollution Control, p. 309-320, 1967.

5. Cleasby, J. L., M. M. Williamson, and E. R. Baumann. Effect of filtration rate changes on quality. JAWWA, 55:7, 869-880, July, 1963.

6. Cleasby, J. L. Filter rate control without rate controllers. JAWWA, 61:4, 181-185, April, 1969.

7. Tuepker, J. Filter performance under varying operating conditions. Proc. of Conf. on Water Filtration. University of Missouri, Rolla. 1965.

8. Hudson, H. E., Jr. Declining rate filtration. JAWWA, 51:11, 1455, November, 1959.

9. Fair, Gordon, J. Geyer, and D. Okun. Water Supply and Wastewater Removal. John Wiley and Sons, Inc., 1966. p. 5-23.

10. Weber, Walter, C. B. Hopkins, and R. Bloom, Jr. Physicochemical treatment of wastewater. Journal, WPCF, 42:1, 83-99, January, 1970.

11. Ives, K. J. Advances in deep bed filtration. Transactions of the Institution of Chemical Engineers, 48:3, T94-T100, April, 1970.

12. Lynam, Bart T. and V. W. Bacon. Filtration and microstraining of secondary effluent. Water Quality Improvement by Physical and Chemical Processes. Water Resources Symposium No. 3., Univ. of Texas Press, Austin, 1970, p. 132-148.

13. Lynam, Bart T., G. Ettelt, and T. McAloon. Tertiary treatment at metro Chicago by means of rapid filtration and micro-strainers. Journal, WPCF, 41, 247, February, 1969. (Hanover)

REFERENCES (cont'd.)

14. Oakley, H. Roy and T. Cripps. British practice in tertiary treatment of wastewater. Journal, WPCF, 41:36, January, 1969.

15. FWPCA. Summary Report, Advanced Waste Treatment, July 1964-July 1967, WPC Research Series, U.S. Dept. of the Interior, Publication WP-20-AWTR-19, p. 21.

16. Smith, Robert. A compilation of cost information for conventional and advanced wastewater treatment plants and processes. Cincinnati Water Research Laboratory, U.S. Dept. of Interior, FWPCA, Advanced Waste Treatment Branch. December, 1967.

Report

to

North Atlantic Division
Corps of Engineers
U.S. Army

on

ULTIMATE DISPOSAL

by

Joseph F. Malina, Jr., Ph.D.
The University of Texas
Austin, Texas

May, 1971

ULTIMATE DISPOSAL

1. Introduction

The ultimate disposal of the solids removed from the wastewater or generated during the treatment of the wastewater, as well as the brines which are produced by the tertiary treatment processes, involves the handling and treatment of these material in such a way that no other environmental problems result. The alternate methods of ultimate disposal include subsurface storage in impervious strata, storage of material in lagoons, or spreading the material on the ground to permit evaporation of the water, ocean disposal, and conversion of the waste to innocuous end-products. This discussion will be limited to the incineration of the sludges produced during treatment, the anaerobic digestion of the sludge prior to disposal of the remaining solids on the land, and the disposal of the brine in the ocean. The other alternates were eliminated because of the hydrological and geological conditions in the Merrimack River Basin.

2. Concepts and Theory

2-1 Incineration

The purpose of incineration is to reduce the volume of solids to be handled by combustion of the volatile solids leaving a solid end-product. The quantity of dry solids and the daily variations in the quantity and composition of the solids will affect the process performance. The concentration of solids in the sludge and the volatile solids content and the heat value of the solids will affect the design and performance of the

incinerator. Low boiling point compounds which tend to be corrosive or require special air pollution control equipment for removal may require special handling. The wide variation in the composition and characteristics of municipal wastewater sludges makes it essential that the incinerator be able to effectively process the sludges over this wide range of variation.

The incineration process may be divided into two distinct phases:

a. drying, and

b. combustion

The drying process involves raising the temperature of the filter cake from the ambient temperature to 212°F, evaporating the water from the solids, and increasing the temperature of the water vapor to the temperature of the exit gases from the incinerator. The heat required to evaporate one pound of water when the sludge entering the incinerator is at a temperature of 60°F and the exit gas temperature is 600°F, is approximately 1300 BTU per pound of water. The heat required for drying the sludge may be derived from the combustion of the volatile solids in the sludge depending upon the amount of moisture which must be removed.

The combustion process, in addition to the fuel and air, requires time, temperature, and turbulence for complete reaction. The temperature in the combustion zone must be sufficiently high to permit ignition of the fuel and to destroy any odor compounds which may be released from the sludge. A minimum temperature of approximately 1400°F is required for combustion of wastewater sludges without the production of odors.

The volatile solids in the sludge provide a source of fuel for the incineration process and the amount of auxiliary fuel required for the incineration of wastewater sludges can be determined from a simple heat balance. Some of the factors to be included in the balance are summarized below. The heat value of sludges can be determined by colorimetric techniques. However, if analytical facilities are not available, the heat value of the sludge can be determined by applying the following equation:

$$Q = a \left(\frac{100\,X_v - b}{100 - X_c}\right)\left(\frac{100 - X_c}{100}\right)$$

in which

Q	=	heat value (BTU/pound dry solids)
X_v	=	volatile solids (percent)
X_c	=	conditioning chemicals (weight percent)
a, b	=	coefficient which characterize the solids, e.g. a = 131 and b = 10 for primary solids in municipal wastewater (fresh and digested). a = 107 and b = 5 for activated sludge

If the chemical composition of the sludge is available, the heat value may also be determined using a modification of the Dulong formula as follows:

$$Q = 14{,}600 + 62{,}000\,(H - \frac{O}{8})$$

in which

Q	=	heat value (BTU/pound dry solids)
C	=	carbon content (weight percent)

H = hydrogen content (weight percent)

O = oxygen content (weight percent)

Typical heat values for various sludges are summarized in Table 1.

Heat is required to evaporate the water from the sludge, to heat the solids to ignition temperature, to heat the incoming air, dehydrate inorganic carbonates, bicarbonates, hydroxides and hydrates in the sludge and to compensate for heat losses.

The heat required to evaporate the water from the sludge was discussed above under drying. The heat available may be calculated using the equation:

$$Q_{in} = \text{(pounds dry solids/hr) (volatile content) (heat value)} = \text{BTU/hr}$$

The heat required to evaporate the water associated with the sludge may be calculated as:

$$Q_{evap.} = \text{(pounds filter cake/hr - pounds dry solids/hr) (BTU/pound)} = \text{BTU/hr}$$

The amount of auxiliary fuel required can be determined from a heat balance for the incinerator.

$$F = \frac{Q_{evap.} + Q_{loss} - Q_{in}}{\text{Heat value of fuel (BTU/lb)}}$$

TABLE 1

HEAT-VALUES OF SLUDGES AND FUELS
(AFTER OWEN, 1957)

Fuel	Combustibles (percent)	Heat-Value (BTU/lb)
Fine screenings	86.4	8,990
Grit	33.2	4,000
Grease and scum	88.5	16,750
Fresh sewage solids	74.0	10,285
Primary sewage sludge	--	7,820
Activated sewage sludge	--	6,540
Digested sludge	59.6	5,290
Digested sewage and garbage solids	49.6	8,020
Waste slufite liquor solids	--	7,900
Semi-chemical pulp solids	--	5,812
Lignite	89.0	10,850
Wood	89.5	8,675
Newsprint	91.8	7,825
Coal	--	10,000 - 14,000
Petroleum	--	19,000 - 20,000

The products of incineration of municipal sludges include an inert residue which is sterile and stack gases, the composition of which will depend to a great extent on the type of installed air cleaning devices. The information available which describes the characteristics of the residue and the stack gas analyses are sparse. The data presented in Tables 2 and 3 indicate that the ash is essentially composed of inorganic materials some of which are water soluble. The results of mass spectrographic analyses of stack gases from an incinerator burning municipal wastewater sludges at 1400°F indicate the following composition:

carbon dioxide	-	8.6 percent
oxygen	-	8.3 percent
nitrogen	-	82.4 percent

Hydrocarbons, oxides of nitrogen, hydrogen sulfide and oxides of sulfur were not detectable in the stack gas. Additional information dealing with the composition of the residues from incinerators and of the stack gases from other installations is required to effectively characterize these components.

The incineration processes available for the burning of sludges include multiple hearth furnaces, rotary kiln furnaces, and fluidized bed reactors. Wet oxidation processes in which the volatile solids in the sludge is oxidized at a temperature of between 350°F and 500°F, and at a pressure of between 1000 and 2000 psi are also available. The degree of oxidization achieved depends upon the sludge, the temperature, pressure, and detention time within the reactor. Typical characterisitcs of the

TABLE 2

Analysis of Ash

Si	20.0%	Ti	0.58%
Ca	15.0	Ba	0.084
Al	5.3	Cr	0.055
Fe	2.7	Pb	0.053
Na	1.4	Mn	0.039
K	1.4	Sn	0.034
Mg	1.3	Other Cations	< 0.01

TABLE 3

Solid Matter Suspended in the Gases

Soluble in H_2O	16.6 percent weight
Soluble in HCl	71.1 percent weight
Unsoluble in HCl	12.3 percent weight
$Fe_2O_3 + Al_2O_3$	30.3 percent weight
$Ca(OH)_2$	39.3 percent weight
Quartz and Silicates	11.7 percent weight
Loss on Ignition (1112°F)	0.6 percent weight

oxidized sludges are shown in Table 4. The amount of organic material in the filtrate increases as the percent oxidation of the sludge increases. The filtrate, therefore, requires additional treatment prior to discharge to the environment.

2-2 Anaerobic Digestion

Anaerobic digestion of wastewater sludges involves the biological degradation of a portion of the organic material in the sludge resulting in a reduction in the volume of the sludge. The destruction of pathogenic organisms and the production of usable products such as a combustible gas and a stable innocuous sludge which can be used as a soil conditioner are also derived from anaerobic digestion process. The required degree of anaerobic digestion depends on the means of ultimate disposal of the sludge. Drying of the sludge on sand beds or applying the sludge to the land would require the greatest degree of stabilization of the organic solids.

In the anaerobic digestion process, there is an orderly and controlled degradation of organic material by bacteria. The energy and building material released are made available to the bacteria and a portion of the organic material is converted to methane and carbon dioxide. The microbial populations are composed of facultative and anaerobic bacteria. The facultative organisms can also use, and in fact, prefer to use molecular oxygen during metabolism; therefore, these microbes protect strict anaerobic bacteria from any free dissolved oxygen which may

TYPICAL CHARACTERISTICS OF PRIMARY-ACTIVATED SEWAGE SLUDGE
of 3.5% Solids and 43 g/l C.O.D. Before and After Wet Air Oxidation

	Before Oxidation	After Low Oxidation			After Intermediate Oxidation			After High Oxidation		
	Raw Sludge	Slurry	Filtrate	Filter Cake	Slurry	Filtrate	Filter Cake	Slurry	Filtrate	Filter Cake
% Cod Reduction	0	5-15	-	-	40-50	-	-	70-80	-	-
% Insoluble Volatile Solids Reduction	0	30-50	-	-	70-80	-	-	92-98	-	-
Filtration Resistance (sec ^{2}/g x 10^7)	3500	6	-	-	6	-	-	10	-	-
Volume, Liters	1.00	*1.05	1.00	0.05	1.00	0.97	0.03	1.00	0.97	0.03
Total Solids, g.	35.1	29.8	8.2	21.6	21.6	7.1	14.5	16.0	5.5	10.5
Volatile Solids, g.	23.4	18.1	5.2	12.9	9.9	4.1	5.8	4.3	3.1	1.2
Ash, g.	11.7	11.7	3.0	8.7	11.7	3.0	8.7	11.7	2.4	9.3
pH	6.0	5.2	-	-	4.9	-	-	6.2	-	-
Moisture in Filter Cake, %	-	-	-	64	-	-	58	-	-	50
Drained Cake Dry Wt. (lb/ft^3) Packed	-	-	-	20	-	-	35	-	-	55
4 hr. Settled Volume, %	100	37	-	-	15	-	-	7.5	-	-
Phosphorus As P**, g.	0.61	0.61	0.18	0.43	0.61	0.06	0.55	0.61	0.02	0.59
Total Nitrogen, g.	1.49	1.49	1.15	0.34	1.49	1.42	0.07	1.49	1.44	0.05
Ammonia Nitrogen, g.	0.57	0.75	0.72	0.03	1.00	0.98	0.02	1.20	1.19	0.01
Total sulfur, g.	0.30	0.30	0.20	0.10	0.30	0.25	0.05	0.30	0.28	0.02
SO_4 as S, g.	-	0.17	0.14	0.03	0.03	0.25	0.05	0.30	0.28	0.02

*Increased Volume Due to Steam Injection

**x 3.065 = PO_4 Content

be introduced into the system with the feed sludge. The anaerobic degradation of solids is a sequential process in which the particulate material is converted to dissolved compounds, primarily organic acids which are then used by the methane forming bacteria to produce methane and carbon dioxide gas. These processes of hydrolysis and gasification take place simultaneously in a well buffered actively digestion system. The composition of wastewater sludges are presented in Table 5 and the environmental conditions required for anaerobic digestion of sludge are summarized in Table 6.

The optimum conditions required for effective anaerobic digestion include thorough mixing of the contents in the digestion tank either mechanically or by means of gas recirculation and maintaining a temperature of between 86°F and 95°F. Under these conditions, the recommended loading to the digestion system is 200 pounds of volatile solids per 1000 cubic feet per day, although higher loadings can be effectively treated if more concentrated sludge was fed to the digester. A detention time of ten days should be sufficient although 15-day detention time is preferable. Improved digestion performance is possible by maintaining the volatile solids and hydraulic loadings as uniform as possible. The characteristics of the sludge leaving an anaerobic digestion system are summarized in Tables 7, 8, and 9.

2-3 Brine Disposal

The brine resulting from the tertiary treatment of municipal wastewater can generally be disposed of in the ocean for the coastal cities.

TABLE 7

Properties of Digester Supernatant
(Municipal Wastewater Sludge)

Item	Standard Rate	High Rate
Temperature ^{o}F	85-90	110-125
Total Solids (mg/l)	4000-5000	10,000-14,000
Total Susp. Solids (mg/l)	2000-3000	4000-6000
BOD (mg/l)	2000-3500	6000-9000
Volatile Solids (mg/l)	650-3000	2400-3800
Alkalinity (MO) (mg/l)	1000-2400	1900-2700
Color (Co-Pt) (mg/l)	3000-4000	4900-6700
H_2S (mg/l)	70-90	190-440
NH_3-Nitrogen (mg/l)	240-560	560-620
pH	7.0-7.6	6.4-7.2
Odor	Slightly Offensive	Offensive
Digester Loadings	0.15 #BOD_5/cu ft/day	0.4 #BOD_5/cu ft/day

TABLE 8

Analysis of Supernatant from Several Plants

Item	Geneva Ill. 1944 (19)	Military Installation 1944 (19)	Downers Grove, Ill. 1963 (20)	Aurora Colorado 1963 (20)	Fort Worth, Texas 1963 (22)	Hogsmill Valley, Eng. 1966 (24)
BOD	542-2852	173-1825	540-4300			
Average	1328	398	2460	2300		
COD			2300-20,000			
Average			11,560			
IOD			80-250			
Average			145			
Total Solids		1320-14,460				1000-4000
Average		3108		14,673	12,100	3300
Suspended Solids	1228-15,484	404-4032	2700-28,980			
Average	5300	1451	5300	11,100		
Alkalinity						500-3100
Average					2630	1900
Nitrogen						
NH_3					510	320
organic					465	60

TABLE 9

	Raw Sludge	Processed Sludge Total	Processed Sludge Liquid	Processed Sludge Solids
Anaerobic Digestion				
Gallons	1000.0	1000.0	754.2	245.8
Solids (lb)	275.0	217.0	12.0	205.0
Nitrogen (lb)	14.0	14.0	3.4	10.6
Phosphorus (lb)	8.0	8.0	0.6	7.4
Wet Air Oxidation				
Gallons	1000.0	1000.0	954.8	45.2
Solids (lb)	275.0	175.0	24.1	150.9
Nitrogen (lb)	10.4	10.4	10.4	0.0
Phosphorus (lb)	2.7	2.7	0.1	2.6

For inland installations, it might be necessary to evaporate the water from the brine or to dispose of the brine in an injection well. The brines resulting from tertiary treatment plants generally will include inorganic materials which will have no adverse effect on the ocean environment.

3. Applicability of the Processes to the System in Question

Incineration and land disposal are the two alternates available for the disposal of the wastewater solids. Anaerobic digestion prior to land disposal provides one possible means of disposing primary and secondary sludges. A portion of the digested sludge could be applied to public lands. However, the hydrological and the geological data indicate that the proximity of the ground water to the ground surface and the highly permeable soils are conditions which would make massive application of digested sludge to the land highly undesirable.

The ultimate disposal of the primary solids, the waste activated sludge and the solids resulting from coagulation in the tertiary treatment scheme can be effectively disposed of by incineration. Dry combustion is preferred over wet air oxidation since in the latter portion the solids are oxidized and the filtrate contains soluble organic and inorganic material as indicated by the data presented in Tables 4 and 9. At the present time, the incinerator would either be a multiple hearth furnace, a rotary kiln furnace, or a fluidized bed reactor.

The solids resulting from the physicochemical treatment plant would also be incinerated. The ash from either of the incineration systems can be reused as an aggregate in building block manufacturing or a portion of the residue can be recycled for sludge conditioning purposes. The residue can also be hauled out to sea for disposal since it contains essentially inorganic salts, a portion of which are water soluble. Disposal of the residue on the land is possible; however, some of the water soluble constituents may be transported to the ground water supply and result in an increase in the total dissolved solids. It is possible that the ion exchange capacity of the soil is such that any soluble cations will be sorbed by the soil.

The scheme for disposal of the brines resulting from the tertiary treatment processes or from the physicochemical treatment scheme will be concentrated by reverse osmosis process and the concentrated stream discharged into the ocean. This system provides for recovery of additional water associated with the spent regenerant from the ion exchange process. The backwash water from filtration can be recycled to either the primary sedimentation basin or the coagulation basin.

4. Pretreatment Requirements

4-1 Incineration

Pretreatment of the sludges is required to reduce the moisture content and to minimize the quantity of auxiliary fuel required for complete combustion of the solids. Alternate processes for sludge

concentration and dewatering are illustrated in Figure 1. For the proposed system, the pretreatment will include thickening of the sludge and dewatering of the thickened sludge on vacuum filters. By concentrating the sludge to a solids content of approximately 35 percent, the sludge should provide sufficient fuel to sustain combustion without any additional auxiliary fuel.

A thickener will be installed prior to the anaerobic digestion system to permit operation at a higher organic loading and reduce the size of digestion tanks required.

The effluent liquid from the thickener and the filtrate from the vacuum filter contain suspended solids and dissolved organic material. These liquids require additional treatment and must be returned to the primary sedimentation basin or to the chemical coagulation system.

5. Experience on the Process to Date

5-1 Incineration

Numerous multiple hearth furnaces and a fewer number of fluidized bed reactors have been used for the incineration of municipal sludges. The consensus of operating experience indicates that with proper maintenance and control, the incinerators can function for long periods of time without any undo and prolonged shutdowns. Sludge burning incinerators have been operated at Minneapolis-St. Paul, Minnesota, Cleveland, Ohio, and Detroit, Michigan, with a minimum of down time.

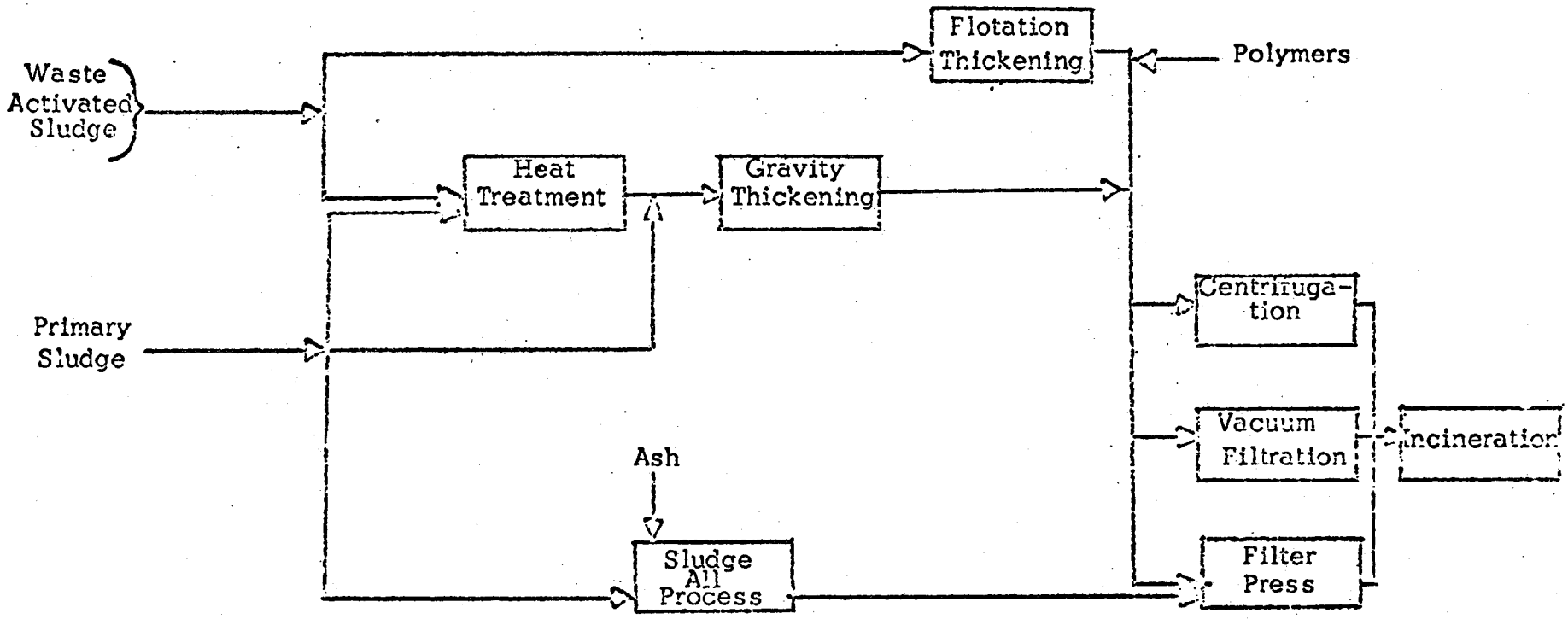


FIGURE 1. ALTERNATE SLUDGE CONCENTRATING PROCESSES

The Detroit plant has six multiple hearth furnaces, each containing 12 hearths. One incinerator was taken out of service each summer for routine maintenance and repairs. Auxiliary fuel was needed only during startup and in those cases when the mositure content of the filter cake exceeded the design value of 65 percent. Burning grease and skimmings in the incinerators had caused problems with the steel shell of the incinerator and with corosion of the scrubbers. When the grease and skimmings were excluded from the incinerator, the corosion problems were significantly less. The auxiliary fuel of this plant included oil and the gas from an anaerobic digester. This dual fuel system resulted in the need for complicated burners and some difficulty was encountered with the automatic control systems. The dry ash handling system was not very effective in maintaining a dust free operation, and required the attention of one or more operators continuously. A water quenched ash handling system would reduce the dust problems and minimize the need for continuous operation.

A multiple hearth furnace is operated at the Nut Island Treatment Plant for disposal of screening and grit. The fuel supply includes gas from the anaerobic digestion system backed up by diesel fuel. The operator indicates a minimum of difficulties were encountered in operation of the incinerator. However, problems with pumping the screenings and the grit had been encountered.

The operation and maintenance of a multiple hearth incinerator at the Bay City Sewage Treatment Plant in Michigan has been carried out for over an eight-year period without excessive problems. This plant is operated only four to six days every two weeks and after an operating period of 96 to 144 hours, the plant is shut down. This sporadic operation of constant warm up and cooling down has not caused any undesirable affects. There are some smoke prlblems, but it is not caused from a defective incinerator, but from a defective stack. The operator feels that good service has been received from this particular incinerator installation.

Multiple hearth furnaces have been successfully operated at over-loads of 25 to 50 percent of the design load although this practice is not recommended. The furnaces should be designed to handle the maximum anticipated load of filter cake otherwise the sludge will accumulate and may cause problems with odor or difficulties with dewatering. The use of plant effluent in scrubbers for fly ash control has resulted in some difficulties with encrustation of the nozzles resulting in the release of odors and particulates in the stack gases.

Synthetic polymers are used to improve the dewaterability of the sludges and result in better filter yeilds; however, these compounds seem to bind more moisture in the sludge cake causing incineration problems. The water seems to be held by the sludge and not easily driven off in the upper parts of the multiple hearth furnace. In these cases, partially burned sludge particles appear in the residue from the incinerator.

Evaluation of the system at the Metropolitan Denver Sewage Disposal District Number One Waste Treatment Plant indicates a number of problems with the operation of the incinerator. The down time of eight days out of the 36-day operating cycle seems to be extremely high compared to the results reported by other installations. Sulfur dioxide and nitrogen dioxide in stack gases were reported. The results of analyses of stack gases from two incinerators indicate that in one case, the particulate matter averaged about 0.038 grains per standard cubic foot of air and that the sulfur dioxide concentration was about 186 ppm and the nitrogen dioxide content averaged 7 ppm. In the other installation, the grain loading was 0.054 grains per standard cubic foot and there was no sulfur dioxide or nitrogen dioxide detected. The operating conditions of the incinerator and scrubber apparently markedly affect the characteristics of the stack gases.

5-2 Anaerobic Digestion

The conventional method of handling and treating the sludges from municipal treatment plants has been anaerobic digestion. Therefore, the literature is replete with reports of successful operations of anaerobic digestion systems. The city of Chicago operates an anaerobic digestion system for handling sludge and the digested sludge is applied to the land to serve as a source of water and soil conditioning material for growing crops. The city of New York has successfully operated anaerobic digestion systems and the residue after digestion is disposed of at sea. The anaerobically digested sludge from the Hyperion Plant in Los Angeles is piped to the ocean.

The anaerobic digestion tanks must be heated to a temperature of 86°F to 95°F for effective digestion. The digestion gas can be used as a source of fuel for heating the sludge, or for the incineration systems.

The anaerobic process can be upset by the introduction of high concentrations of toxic materials such as heavy metal, sulfides and other materials which might have a deleterious effect on the methane forming bacteria. Anaerobically digested sludge does not dewater as readily as raw sludge; therefore, the cost of dewatering the digested sludge would be higher than that for raw sludge.

Digesting the sludge anaerobically prior to incineration does not seem feasible since in the anaerobic digestion process, a portion of the volatile solids are destroyed. These volatile solids also serve as a source of fuel in the incineration process. Since the anaerobically digested sludges are more difficult to dewater, the amount of moisture which must be evaporated in the incinerator also increases. The methane produced during anaerobic digestion can in part compensate for some of the loss of the fuel and increased fuel requirements. However, if the ultimate disposal of solids is incineration, there does not seem to be any real advantage to undergoing anaerobic digestion prior to incineration.

The very permeable soil and close proximity of the ground water table to the surface are not conditions which are favorable for large-scale land disposal of digested sludge. A portion of the sludge could

possibly be digested and applied to public lands and thoroughfares to provide the moisture and some fertilizer. Therefore, land disposal of digested sludge does not seem feasible.

6. Sludge Handling and Disposal

Sludge production and disposal will be evaluated for the following four areas and conditions:

a. Boston

b. Lowel-Lawrence-Haverhill

c. Lowell-Lawrence-Haverhill will base level treatment of industrial wastes

d. Lake Winnipesaukee

Two wastewater treatment schemes will be used as the basis for estimating the composition and quantity of sludge. These systems are:

a. primary treatment, followed by activated sludge and tertiary treatment

b. physicochemical treatment

Boston

Sludge Production

Preliminary Treatment

Grit 4.0 cu ft/million gallons containing 50% moisture, 50% volatile solids, screenings use 1.0 cu ft/million gallons. Therefore, screenings collected are 424 cu ft/day; and 424 (4.0) = 1696 cu ft of grit.

Assume grit has specific gravity of 2.0. The quantity of grit is 1696 (2) (62.4) = 212,000 lb/day, or 106 T/day grit.

Primary Treatment

Efficiency

BOD removal 35%

SS removal 70%

Solids Removed

424 (260)(8.34)(0.7) = 644,000 lb/day Dry Solids at 5% solids

Sludge Volume

$$\frac{644{,}000 \text{ lb}}{(0.05)\text{lb/lb}} \frac{\text{cu ft}}{62.4\text{lb}} = 206{,}000 \text{ cu ft/day} \qquad 1{,}541{,}000 \text{ gal/day}$$

Secondary Treatment

BOD removed = 0.65 (175) - 25 = 88.8 mg/l

Sludge Produced

$\Delta X = a\, S_r - b\, X$ based on $t = \frac{24\ (0.65)\ 175}{2000\ \ (0.3)} = 4.5$ hr or

0.188 days

$V = 1.25\ (424)(0.188) = 99.6$ million gallons

$\Delta X = 0.73\ (88.8)(424)\ 8.34 - 0.075\ (99.6)(2000)(8.34)$

= 229,000 - 125,000

= 104,000 lb/day Volatile Solids

or 130,000 lb/day Total Solids

65 tons/day at 1% solids

Sludge Volume

$$\frac{130,000}{0.01\ (62.4)} = 208,300 \text{ cu ft/day} \qquad 1,555,000 \text{ gal/day}$$

Coagulation

500 lb solids/million gallons

solids = 500 (424) = 212,000 lb/day

@ 1.5% solids

$$\text{Sludge Volume} = \frac{212,000}{0.015\ (62.4)} = 226,500 \frac{\text{cu ft}}{\text{day}}$$

or 1,695,000 gal/day

Summary of Sludge Produced in Boston

Type	lb/day	solids content	volatile content	volume
Grit	212,000	50%	50%	1.696 cu ft/day
Primary Sludge	644,000	5%	65%	1.541 MGD
Secondary Sludge	130,000	1%	80%	1.555 MGD
Chemical Sludge	226,500	1.5%	40%	1.695 MGD
	1,000,500*			4.791 MGD

*Excluding grit

Concentration of solids after mixing three sludges (excluding grit) $\frac{100,500}{4.791\ (8.34)} = 25,000$ mg/l or 2.5%

This sludge can be thickened to 6% solids @ mass loading of 20-25 lb/sq ft day

Incineration

The volatile solids content of the sludge is about 60% The vacuum filter cake will contain 30% solids at a loading of about five to ten lb/sq ft-hr. Assume 2% by weight polymer required

∴ Solids Added = 20,000 lb/day

Design Example

Design Data

Dry Solids (including polymers and grit)	1,232,500 lb/day
Per Cent Volatile Sludges	60%
Filter Cake Moisture	70%
Heat Value of Solids	10,000 BTU/lb volatile
Supplementary Fuel	Natural Gas
Heat Value of Fuel	1,000 BTU/cu ft
Operating Schedule	24 hours/day 7 days/week

Dry Solids per Hour

1,232,500/24 = 51,350 lb/hr

$$\text{Filter Cake} = \frac{1{,}000{,}500 \text{ lb/day}}{0.3 \frac{\text{lb solids}}{\text{lb cake}} \; 24} = 139{,}000 \text{ lb/hr}$$

$$\text{grit } \frac{212{,}000}{24} \qquad 8{,}850 \text{ lb/hr}$$

147,850

51,350

96,500 lb/hr

Heat Balance

Basis

800°F	out gas temperature
75%	excess air
50%	relative humidity
60%	ambient temperature

Air Requirements

Primary air $30{,}810 \text{ lb/hr} \left(\frac{8 \text{ lb air}}{\text{lb VS}}\right) = 246{,}500 \text{ lb/hr}$

Excess air $(0.75)(246{,}500)$

Total air $\frac{184{,}900}{431{,}400 \text{ lb/hr}}$

Water in Air

$$\frac{0.0055 \text{ lb}}{\text{lb air}} \left(431{,}400 \frac{\text{lb air}}{\text{hr}}\right) = \frac{2373 \text{ lb } H_2O}{\text{hr}}$$

Heat Required	BTU/lb	lb/hr	10^6 BTU/hr
Water in Sludge Cake	1400	96,500	135.10
Water in Air	0.47(800-60)	2,370	0.83
Volatile Solids	4000	30,810	123.25
Ash	130	20,540	2.67
Radiation			14.50
Cooling Air			8.30
			284.65

Heat Available

Volatile Solids $10{,}000 \text{ BTU/lb} \times 30{,}810 \text{ lb/hr} = 308.10 \times 10^6$

The heat available exceeds the heat required; therefore, auxiliary fuel is not necessary. However auxiliary fuel must be provided for startup of the incinerator. Provisions for use of auxiliary fuel should be included in the final design to insure continuous operation in the event the characteristics of the sludge and filter cake change.

Incinerator Size

$$\frac{\text{Heat Available in Sludge}}{\text{Water to be evaporated}} = \frac{308.1 \times 10^6}{96{,}500 \text{ lb/hr}} = 3200 \frac{\text{BTU}}{\text{lb water}}$$

Use a hearth loading of 10.0 filter cake per square foot of hearth area per hour.

$$\text{Area required} = \frac{147{,}850 \text{ lb/hr}}{10 \text{ lb/sq ft-hr}} = 14{,}785 \text{ sq ft}$$

Use furnaces which have an outside diameter of 22 feet 3 inches and 8 hearths with area of 2117 sq ft.

$$\frac{14{,}785 \text{ sq ft}}{2117 \text{ sq ft/unit}} = 6.99 \text{ units}$$

use seven units

Anaerobic Digestion

Total Solids = 1,000,500 lb/day (excluding grit and screenings)

Volatile Solids = 613,200 lb/day

Volume of digesters required

$$\frac{613,200 \text{ lb vs/day}}{0.2 \text{ lb vs/cu ft-day}} = 3,066,000 \text{ cu ft}$$

Volume of sludge = 640,500 cu ft/day at 2.5%

Volume of thickened sludge at 6% = 267,000 cu ft

Detention time in digestion tank

$$\frac{3,066,000}{267,000} = 11.48 \text{ days} \quad \underline{\text{ok}}$$

Use 9 digestion tanks

125 ft in diameter and 30 feet SWD

Volume of each tank is 375,000 cu ft

Total Volume is 3,375,000 cu ft

Available detention time = 12.64 day

Loading = 182 lb vs/1000 cu ft-day

Methane production = 8 cu ft/lb vs added

$$\therefore \frac{8 \text{ cu ft}}{\text{lb vs}} (613,200) = 4,900,000 \text{ cu ft.}$$

Quantity of solids for final disposal = (1 - 0.60)(613,200) +

387,300 volatile solids destroyed Fixed Solids

= 632,580 lb

Assuming 10% solids in the sludge, the volume of digested sludge is V = 102,000 cu ft/day or 758,300 gal/day + Grit

<u>Lake Winnipesaukee</u>

Solids Production

Type	lb/day	solids content	volatile content	volume
Screenings	---	--	--	9.72 gal/day
Grit	4,870	50%	50%	39 cu ft/day
Primary Sludge	8,966	5%	65%	21,500 gal/day
Secondary Sludge	1,960	1%	80%	23,490 gal/day
Chemical Sludge	4,860	1.5%	40%	38,860 gal/day
	15,786*			83,850

(excluding grit

Concentration of mixed sludges excluding grit is 2.26%.

Thickened sludge to 6% solids.

Vacuum filter cake contains 30% solids at 60% volatile solids.

2% chemicals by weight = 316 lb/day.

Total quantity of Solids to Incinerator = 20,976 lb/day

use 21,000 lb/day or 10.5T/day

Dry solids	875 lb/hr
Filter Cake	2,192 lb/hr
Grit	203 lb/hr
Dry Solids	2,395 lb/hr
Water	875
	1,520 lb/hr
Heat Required	4.52×10^6 BTU/hr
Heat Available	4.91×10^6/BTU/hr

$$\frac{\text{Heat Available}}{\text{water}} = 3{,}230 \frac{\text{BTU}}{\text{lb water}}$$

Use hearth loading of $\frac{\text{10 lb filter cake}}{\text{sq ft-hr}}$

Area Required $= \frac{2,395 \text{ lb/hr}}{10} = 23.9$ sq ft

use single furnace

8 ft - 6 in in diameter with 8 hearths with area = 256 sq ft

or

10 ft - 0 inches in diameter with six hearths with area = 276 sq ft

Anaerobic Digestion

Volume required at loading of 0.2 lb vs/cu ft-day = 46,700 cu ft.

Detention time with six percent sludge = 11.10 days

Use two digesters 40 ft in diameter and 20 feet SWD each

with volume = 25,000 cu ft.

Total volume = 50,000 cu ft.

Detention time = 11.9 days

Loading = 187 lb vs/1000 cu ft - day

Sludge for disposal on land

Solids = 10,182 lb/day

at 10% Solids

Volume = 1632 gal/day or 218 cu ft/day + 39 cu ft grit

Methane produced = 74,750 cu ft/day

Lowell-Lawrence-Haverhill (No Ind Treat)

Sludge Production

Type	lb/day	solids content	volatile content	Volume
Screenings	---	--	--	105 cu ft/day
Grit	50,500	50%	50%	420 cu ft/day
Primary Sludge	173,000	5%	65%	0.415 gal/day
Secondary Sludge	50,800	1%	80%	0.610 MGD
Chemical Sludge	60,500	1.5%	40%	0.484 MGD
	284,300*			1.509 MGD

*excluding grit

Solids concentration of mixed sludges = 2.02%

This sludge is concentrated to 6% solids and dewatered to 30% solids in the filter cake.

Polymer addition is 2% by weight = 5700 lb/day

Total Solids = 340,500 lb/day or 170.3 T/day

Filter Cake	39,500 lb/hr
Grit	2,100 lb/hr
Dry Solids	41,600 lb/hr
Water	14,200 lb/hr
	27,400 lb/hr
Heat Required	76,080,000 BTU/hr
Heat Available	75,300,000 BTU/hr

Auxiliary fuel is required

$$\frac{\text{Heat Available}}{\text{water}} = \frac{2,748 \text{ BTU}}{\text{lb } H_2O}$$

Use hearth loading of 8 lb filter cake per sq ft-hr

Area required $= \frac{41,600}{8} = 5200$ sq ft

use 3 furnaces

each 19 ft - 6 inches in diameter with 8 hearths area $= 2060 \frac{\text{sq ft}}{\text{furnace}}$

Total area = 3 (2060) = 6180 sq ft.

Anaerobic Digestion

Volume required at loading of 0.2 lb vs/cu ft-day = 904,000 cu ft.

Detention time = 13.3 days

Use 4 units each 100 ft in diameter with 30 ft SWD each has a volume of 235,000 cu ft

Total volume is 940,000 cu ft

Detention time is 13.8 days

Loading is 192 lb/1000 cu ft-day

Solids remaining = 232,400 lb/day at 10% solids

Volume = 37,250 cu ft/day or 278,500 gal/day

Methane Produced = 1,446,000 cu ft/day

Lowell-Lawrence-Haverhill (with BLT)

Sludge Production

Type	lb/day	solids content	volatile content	Volume
Screenings	---	--	---	105 cu ft/day
Grit	50,500	50%	50%	420 cu ft/day
Primary Sludge	150,000	5%	65%	0.360 MGD
Secondary Sludge	44,350	1%	80%	0.532 MGD
Chemical Sludge	60,000	1.5%	40%	0.480 MGD
	254,350*			1.372 MGD

*Excluding grit

MULTIPLE HEARTH OPERATING DATA AND COSTS

Population equivalent	10,000	20,000	50,000	100,000	1,000,000
Sludge incinerated, tons/wk (wet basis)	28.0	56.0	139.0	278.0	2,780.0
Sludge incinerated tons/wk (dry basis)	7.0	14.0	34.75	69.5	695.0
Operating schedule, hr/wk	35	35	70	70	168
Furnace feed, lb/hr	1,600	3,200	3,960	7,920	33,000
Furnace required	10 ft.-9 in. OD 5 Hearth ..	14 ft.-3 in. OD 5 Hearth	14 ft.-3 in. OD 6 Hearth	18 ft.-9 in. OD 6 Hearth	Two-22 ft.-3 in. OD 8 Hearth
Installed cost, $	120,000	185,000	200,000	310,000	750,000
Weekly fuel cost, $	27.50	45.00	30.00	45.00	50.00
Weekly power cost, $	9.00	12.00	13.00	25.00	165.00
Total utility cost, $	36.50	57.00	43.00	70.00	215.00
Operating cost, $/ton dry solids	5.20	4.06	1.24	1.01	0.31
Filtration cost, $/ton dry solids	8.00	8.00	8.00	8.00	8.00
Maintenance cost, $/ton dry solids	.70	.60	.60	.50	.40
Total disposal cost, $/ton dry solids	13.90	12.66	9.84	9.51	8.71

Cake moisture 75%
Volatile content ..... 65%; 10,000 Btu./lb. volatile solids
Solids removal 90%

EXPECTED INCINERATION COSTS (a)

Plant Flow MGD	Dry Sewage Solids Tons/Day	Required Furnace Size	Capital Cost Installed	Operating Cost, $/Ton D. S. S.			
				Fuel (b)	Power (c)	Maint.	Total
1.0	0.93	10'-9" O.D. x 5 Hearth	$150,000	$6.35	$2.40	$0.50	$9.25
2.0	1.85	14'-3" O.D.	$200,000	5.50	1.64	0.40	7.55
5.0	4.63	14'-3" O.D.	$225,000	2.95	1.15	0.25	4.35
10.0	9.25	16'-9" O.D.	$280,000	2.50	1.00	0.15	3.65
15.0	13.88	18'-9" O.D.	$330,000	1.90	0.75	0.10	2.75
20.0	18.5	22'-3" O.D.	$450,000	2.30	0.65	0.10	3.05

(a) Costs do not include dewatering equipment, Chemicals, or Power

(b) Cost of Natural Gas $0.08/Therm (100,000 BTU)

(c) Cost of Power $0.02/KWHr.

Reference:

Harris, S. M., "Incineration-Multiple Hearth Furnaces", Water & Sewage Works (Aug. 1967), pp 307-312.

REFERENCES

1. Balakrishnan, S., Williamson, D. E., Okey, R. W., "State of the Art Review on Sludge Incineration Practice" for the Federal Water Quality Administration, Department of the Interior. Program # 17070 Div., Contract #14-12-499. (April 1970).

2. Burd, R. S., "A Study of Sludge Handling and Disposal." U. S. Department of the Interior, Federal Water Pollution Control Administration. (May 1968).

3. Corey, Richard C., "Principles and Practices of Incineration." Published by Wiley-Interscience (1969).

4. Ducan, G. J., and Levin P., "Mathematical Model of Sewage Sludge Fluidized Bed Incinerator Capabilities and Costs." For the Advanced Waste Treatment Research Laboratory, Robert A. Taft Water Research Center. U.S. Department of the Interior, Federal Water Pollution Control Administration, (September 1969).

5. Gloyna, Earnest F., and Eckenfelder, W. W., Jr., "Water Quality Improvement by Physical and Chemical Processes" Water Resources Symposium No. 3, Center for Research in Water Resources, The University of Texas Press (1970).

6. Harris, Sydney M., "Incineration-Multiple Hearth Furnaces." Water and Sewage Works, pp. 307-311, (August 1967).

7. Malina, Joseph F., Jr., "Separate Sludge Digestion," Manual of Wastewater Operations, Chapter 18, pp. 319-360 (1971).

Report

to

North Atlantic Division
Corps of Engineers
U.S. Army

on

WASTE TREATMENT IN THE MERRIMACK RIVER BASIN BY ION EXCHANGE

by

R. L. Sanks, Ph.D.
Professor of Civil Engineering and Engineering Mechanics
Montana State University
Bozeman, Montana

May, 1971

1. THE PROBLEM DEFINED

Regardless of other contaminants, the salt content of the Merrimack River is extremely low as given by Valentine (1971) in Table I and its quality in this respect is very high. Use by a municipality adds about 350 mg/l of dissolved solids to the water as shown by Neale (1964) in Table II. Table III shows the average "use increment" added to the average ionic content of the Merrimack River. This would approximate the expected analysis of secondary effluent from a city using the Merrimack River as its water supply.

At the second meeting of the Board of Consultants, 4 May, 1971 at Lafayette, Indiana, it was decided there would be two kinds of waters to be demineralized:

Water A: 275 mg/l TDS demineralized to 220 mg/l TDS
Water B: 500 mg/l TDS demineralized to 100 mg/l TDS.

The higher ionic content of Water B would be the result of removal of ammonia by breakpoint chlorination.

On the basis of dissolved solids the quality of each of these three hypothetical waters (Table III, Water A, and Water B) is good. The hardness of each is very low and all ions are within U.S.P.H.S. standard limits. Except for $PO_4^{\equiv}$, the ionic content of the secondary effluent seems quite acceptable and, since coagulation will reduce the $PO_4^{\equiv}$ to very low values, there is no inherent need to demineralize these waters. However, the Corps of Engineers may be reluctant to condone treatment methods which degrade the water in one respect (dissolved solids content) regardless of the benefits in all other respects. Furthermore, it might be difficult to persuade the public to accept reused water unless it were equal to or better than river water in every respect. Hence there may be a need to demineralize the treated wastewater even though the dissolved solids content would be acceptable elsewhere.

TABLE I. IONS IN MERRIMACK RIVER AT LOWELL
Data obtained from Corps of Engineers, May 1971

Cation	mg/l	me/l[a]	Anion	mg/l	me/l
NH_4	--	--	CO_3	0	0
Na	8.8	0.382	HCO_3	9.3	0.152
K	1.0	0.026	Cl	13.3	0.377
Ca	5.8	0.290	SO_4	11.6	0.241
Mg	1.2	0.099	NO_3	2.8	0.045
Fe	0.06	0.003	F	0.3	0.016
Mn	0.09	0.005	Total	37.3	0.831
Total	16.95	0.805	SiO_2 = 6.1 mg/l (assume colloidal)		

TABLE II. AVERAGE USE INCREMENT IN 22 U.S. CITIES
From Neale (1964)

Cation	mg/l	me/l	Anion	mg/l	me/l
NH_4	15	0.831	CO_3	1	0.033
Na	66	2.870	HCO_3	100	1.640
K	10	0.256	Cl	74	2.086
Ca	18	0.900	SO_4	28	0.583
Mg	6	0.494	NO_3	10	0.161
Fe	--	--	NO_2	1	0.022
Mn	--	--	F	--	--
Total	115	5.351	PO_4	24	0.756
			Total	238	5.281

TABLE III. APPROXIMATE ANALYSIS OF A SECONDARY EFFLUENT
IN THE MERRIMACK RIVER BASIN
Derived from Tables I and II

Cation	mg/l	me/l	%	Anion	mg/l	me/l	%
NH_4	15	0.83	14	HCO_3	109	1.83	30
Na + K	86	3.53	57	Cl	87	2.46	40
Ca	24	1.19	19	SO_4	40	0.82	13
Mg	7	0.59	10	NO_3	13	0.20	4
Total	132	6.14	100	PO_4	24	0.76	13
				Total	273	6.07	100

TDS = 405

[a] me/l = milliquivalents/liter

2. SUMMARY OF CONCEPT AND THEORY

Ions can be exchanged by many materials including (from poor to good): soil, clay, glauconite (green sand), zeolite, sulfonated coal, clinoptilolite, and synthetic resins. The synthetic resins consist of polystyrene - divinylbenzene, phenolic, or acrylic matrices and functional groups (exchange sites) of sulfonic acid (strong acid resins) carboxylic acid (weak acid resins) quaternary ammonium (strong base resins), and primary secondary or tertiary amines (weak base resins). The exchange is reversible, equivalent -- i.e. one me (milliequivalent) of Na exchanges for one me of Ca -- and proceeds toward an equilibrium at a kinetic rate dependent upon many complex interacting phenomena. A strong acid cation exchanger in hydrogen form releases H^+ ion in preference for Na^+.

$$H \cdot R + Na^+ = H^+ + Na \cdot R$$

and a strong base anion exchanger prefers Cl^- over OH^-.

$$R \cdot OH + Cl^- = R \cdot Cl + OH^-$$

So the product is exhausted resin and HOH, water.

Strong acid resins must be regenerated with massive doses of a strong acid such as sulfuric acid, whereas weak acid resins, which prefer H^+ ion over all others, can be regenerated with weak acids or weak concentrations of strong acids. Similarly, strong base resins require caustic for regeneration but weak base resins can be regenerated with almost any base such as lime or ammonia.

The cycle of operation is: exhaustion (the service run), backwash, regeneration, and rinse. A high level of regeneration can result in water equal to or exceeding the purity of distilled water, and most ion exchange plants (and most expertise) is directed toward this purpose. But in partial demineralization, or desalination, it is more important to reduce the waste volume from the process and to minimize the cost than to produce a high quality product.

A. Strong Acid - Weak Base Fixed Bed Process

This is the simplest system and for small plants may well be the least expensive. If the feed contains substantial alkalinity, a degasifier following the strong acid reactor is needed and reduces cost. Using split treatment and practicing the strictest water conservancy in regeneration, backwash, and rinsing, Sanks and Kaufman (1967) operated a bench-scale pilot plant which produced a waste volume from the ion exchanger of only 3.3% of total blended product. A schematic diagram is shown in Fig. 1. Solids reduction from 850 mg/l to 500 mg/l was estimated to cost 19¢/1000 gals including amortization over 11 years but excluding operating labor. However, the estimate was conservatively based on an 11 year life. The cation exchange efficiency, 83%, was later increased by Sanks (1967) to 92% by using 26% weak acid resin in the cation reactor. Were the estimate made on the same basis as other investigators, the cost would have been substantially less.

B. Sul-biSul Process

This is similar to the previous process except the anion reactor is loaded with a strong base resin (Dowex SBR) which operates on a sulfate-bisulfate cycle. Exhaustion is

$$R_2 \cdot SO_4 + HCl = R \cdot HSO_4 + R \cdot Cl$$

The resin can be regenerated with large quantities of water or small quantities of water and lime

$$R \cdot HSO_4 + R \cdot Cl \xrightarrow{Ca(OH)_2} R_2 \cdot SO_4 + HCl$$

The controlling mechanism is equilibrium between HSO_4^- and $SO_4^=$

$$SO_4^= + H^+ = HSO_4^-$$

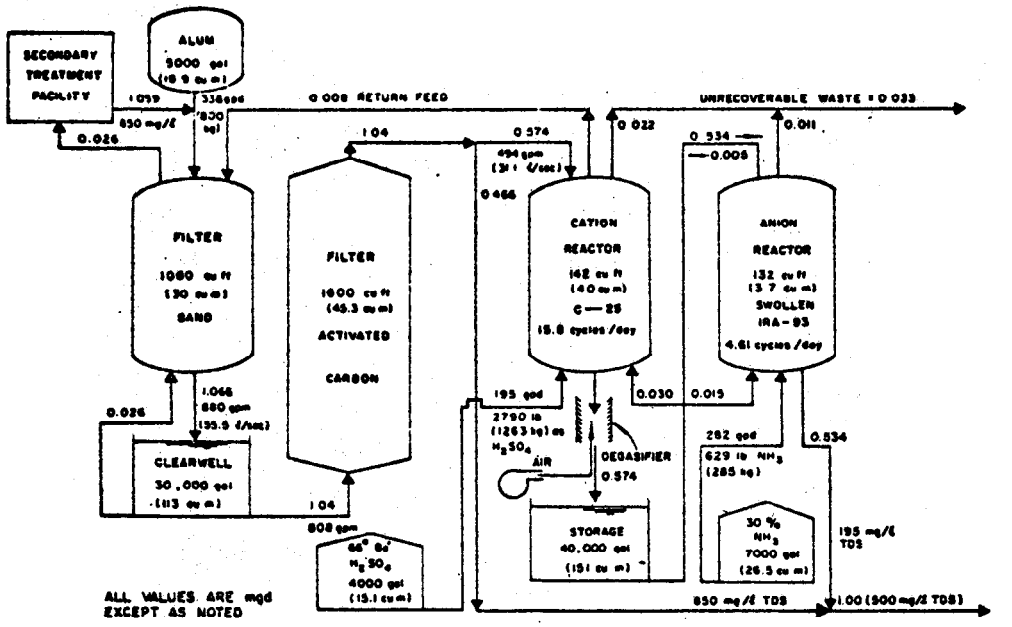


FIG. 1. Schematic flow diagram for 1.00-mgd (3,785-cu m/day) ion-exchange plant to treat 850 mg/l. TDS secondary effluent.

Sanks and Kaufman (1967)

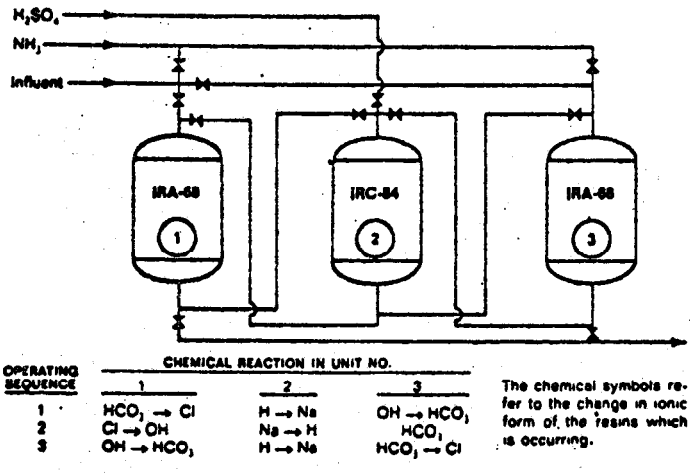


FIG. 2. THREE BED DESAL PROCESS
Kunin (1970)

Lime is the most economical regenerant which accounts for the low cost of the Sul-biSul process. Organic fouling, characteristic of strong base resins regenerated to OH^- form, is absent in the Sul-biSul process. However, the exchange capacity of strong base resin is only half that of weak base resin and furthermore only half of the sites are available for exchange, so the reactor is large. The choice between the Sul-biSul process and the strong acid - weak base system depends strongly upon the water analysis. The Sul-biSul process requires a substantial proportion of $SO_4^=$ in the water.

C. Desal Process

Unlike all other processes, the Desal process does not require any strong electrolyte resins. With its weak electrolyte resins the Desal process is about 90% efficient in both cation and anion exchange. It is particularly effective for high solids (TDS > 500 mg/l) waters.

The process (Fig. 2) works because a certain anion resin, IRA-68, in HCO_3^- form can sorb Cl^-, NO_3^- and $SO_4^=$. The effluent from the first reactor containing Na^+, Ca^{++}, Mg^{++}, and HCO_3^- can be decationized by a weak acid resin (IRC-84) in the second reactor. The third reactor, containing IRA-68 in OH form, releases OH^- and sorbs HCO_3^-. The first reactor is regenerated with caustic, lime, or ammonia, and the second is regenerated with any acid. The reactors are now in their original state except that their positions are reversed from the original form

$$HCO_3^-, H^+, OH$$

to the form

$$OH^-, H^+, HCO_3^-$$

and so the flow, originally from left to right is reversed to right to left.

The Desal process can be modified to a two-reactor system (as in Fig. 3) or to a combination of chemical treatment and ion exchange as in Fig. 4. A complex, closed-system desalination system for complete recycle with no liquid waste, is currently planned for a Montana site.

To work properly, the pressures must be high (above 50 psi) and CO_2 addition may be required. The valving is complicated. It requires careful operation, but with automation these difficulties can be resolved.

D. Continuous Counter-Current Ion Exchange (CCIX)

Chem-Seps. In the Higgins or Chem-Seps process, the resin, held in a closed loop, is pulsed around the loop in short movements at frequent intervals. This keeps the breakthrough curve entirely within the feed chamber (Fig. 5), keeping leakage of ions to a minimum and giving maximum efficiency in regeneration. By using 50% weak acid and 50% strong acid resins in the cation reactor, exchange efficiencies of 90% have been achieved. Equal results have been realized with the anion reactor.

Demineralizing a water from 500 mg/l TDS to 10 mg/l, the Chem-Seps process can operate at 90% efficiency with a total waste stream of only 3% of the influent. The smallest competitive plant would produce 300,000 gpd, so the process is not adapted to small flows.

Asahi-Graver. In this process, shown in Fig. 6, the resin is exhausted in one vessel and transferred to another for regeneration and to still another for washing. As in the Chem-Seps process the resin moves in short pulses.

The advantages claimed are highly efficient operation and less expensive reactors.

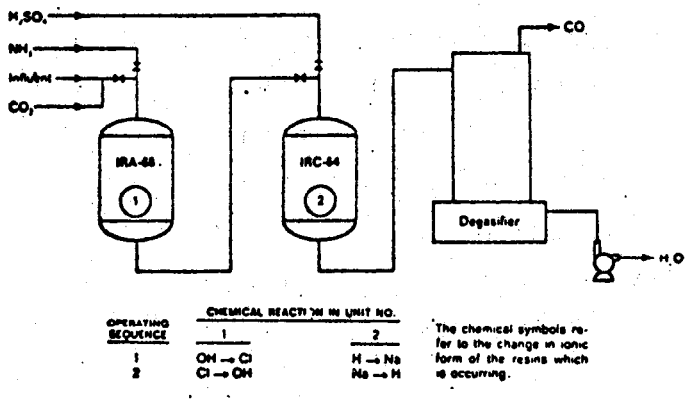


FIG. 3. TWO-BED DESAL PROCESS
Kunin (1970)

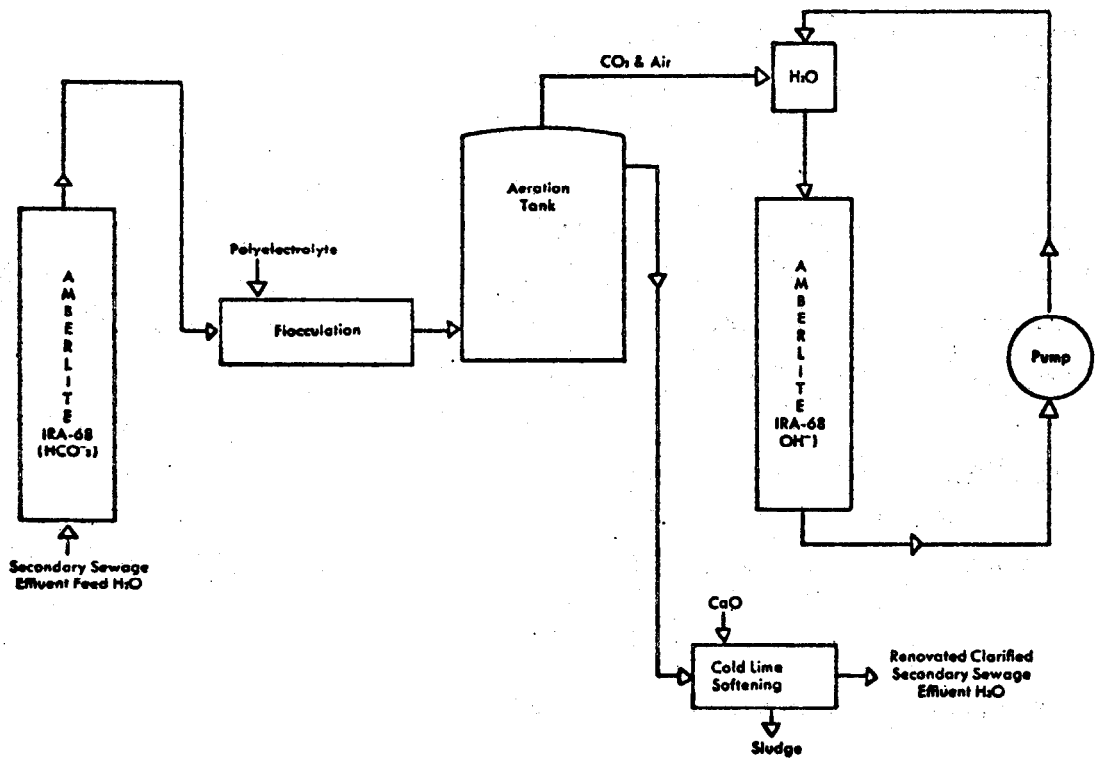


FIG. 4. MODIFIED DESAL PROCESS FOR RENOVATION OF SECONDARY SEWAGE EFFLUENT WITH CO_2 RECOVERY
Kunin (1969)

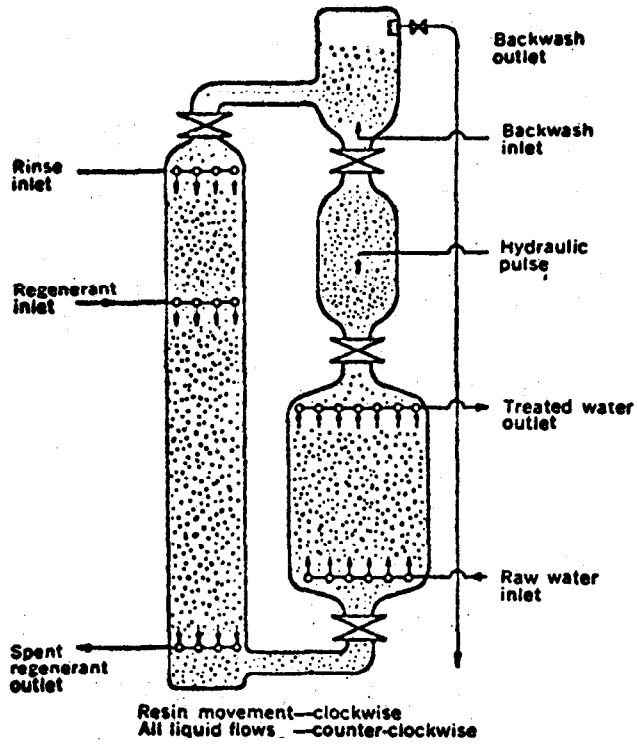


FIG. 5. CONTINUOUS COUNTERCURRENT CONTACTOR FOR CHEM-SEPS PROCESS
Kunin (1970)

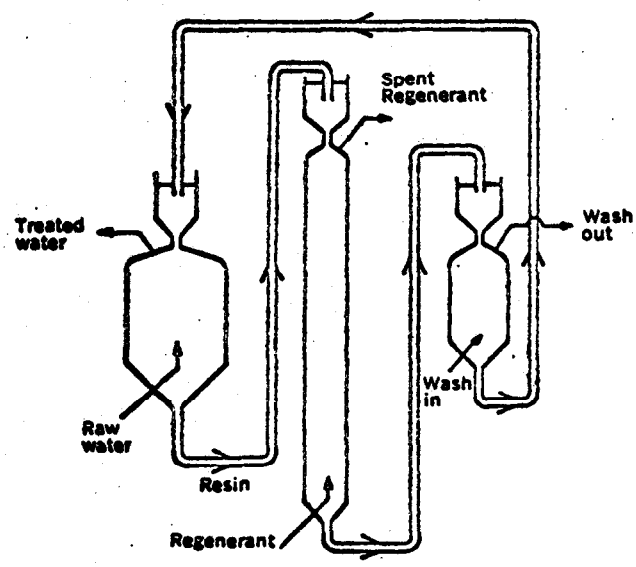


FIG. 6. CONTINUOUS COUNTERCURRENT CONTACTOR FOR ASAHI-GRAVER SYSTEM
Kunin (1970)

Progressive Mode. The Permutit Company has abandoned CCIX (saying it is too difficult to control) in favor of a highly modified multiple fixed-bed system. Each reactor is regenerated, rinsed, and placed into service in turn and the feed is routed through several reactors in series as shown schematically in Figure 7. Only completely exhausted resins are regenerated, so the efficiency is high, perhaps as much as 95%. The final tank through which feed passes is freshly regenerated so the leakage is low.

In contrast to ordinary fixed-bed reactors in which flow rates ordinarily do not exceed 10 gpm, Progressive Mode uses 40 gpm per sq ft.

L*A Type CF System. The L*A Water Conditioning Company packs resin tightly into zones subject to upflow service and downflow regeneration. They use a cyclic staging arrangement whereby tanks instead of resins are artificially moved to obtain continuous operation.

E. Selective Ion Exchange

While there are some chelating resins that are highly selective for certain ions, there seem to be no exchangers highly selective for Cl^-, NO_3^- or Na^+. However, there is an exchanger highly selective for ammonia. Hector clinoptilolite (a natural zeolite) is selective for NH_4 in the presence of Na^+, Ca^{++}, and Mg^{++} ions. One cubic foot is capable of removing 97% of the ammonia from 2000 gallons of secondary effluent. It is regenerated with lime. Ammonia can be removed from the waste regenerant in an air stripping tower and exhausted to the atmosphere or recovered in a stream stripper for regenerating anion exchangers, for fertilizer or for other uses. The lime can be reused with small amounts of make-up lime. Nitrate specific exchangers have been developed.

Summary. There are many processes, each with shortcomings and advantages that shift with changing flows, water analyses, and objectives. The proprietary ones have persuasive adherents and it will require

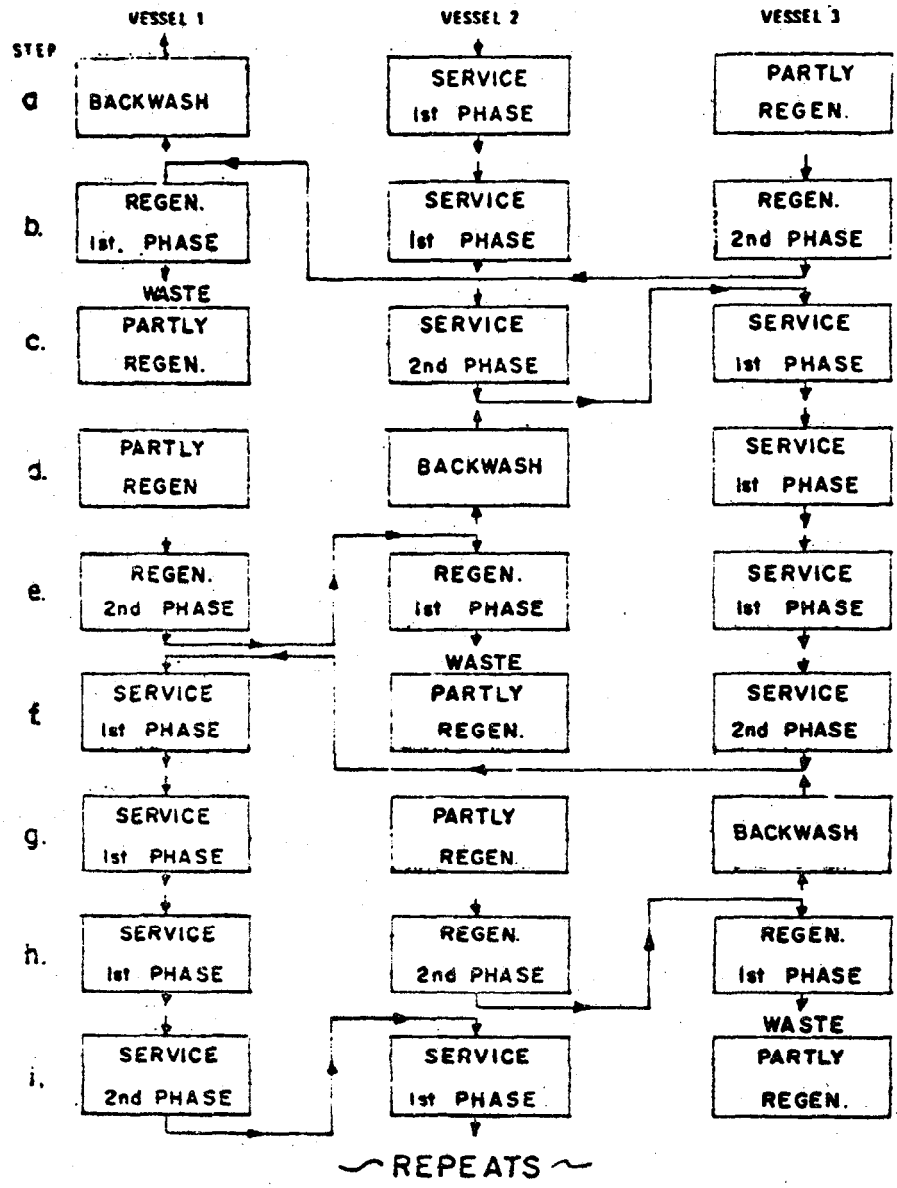


FIG. 7. STEP SEQUENCE FOR THREE-BED PROGRESSIVE-MODE ION EXCHANGE
Anderson and Voss (1970)

considerable expertise to select the most appropriate. The cost of tertiary treatment is great enough so that pilot plant studies should always precede full-scale installation.

3. APPLICABILITY

Ion exchange is readily applicable to either of the two schemes proposed by the Board of Consultants. The process can be used to demineralize part of the waste stream which is then blended with by-passed water. Or all of the water can be passed through the exchanger which can be operated to achieve any degree of solids removal. One of the advantages of ion exchange is its flexibility.

Unfortunately, the selectivity of ions cannot be varied at will. Ammonia and sodium break through before calcium or magnesium, and chloride breaks through before sulfate. As described in Item 2E, one exception of note is clinoptilolite. There are some chelating resins that are highly selective for certain ions, so there may be hope that a resin selective for chloride may be found.

4. DESIGN CRITERIA

These will differ depending upon the process selected. Criteria for two processes are given herein.

A. Strong Acid - Weak Base Fixed Bed

A service flow rate of 4 gpm/cu ft or 10 gpm/sq ft (for 2½ ft beds) can be assumed. Cation beds can operate in service flow 70% of the time and anion beds 85% of the time. Overall exchange efficiency can be roughly estimated at 87% and leakage (of ions leaving the system) can be estimated at 20%.

B. Chem-Seps

The cost curve in Fig. 8 is for a complete plant and housing and includes all facilities except outside piping. Capital and operating costs (except for regenerant) given under Item 8, were obtained from the manufacturer and are assumed to be accurate. Regenerating costs are given in Item 9.

C. Labor

Labor is an unknown factor and an important one, especially for small plants. Ion exchange plants operate automatically and require perhaps only one or two hours attention per day. If political factors govern, it may be necessary to provide 24-hr/day surveillance, which would greatly increase labor costs.

5. PRETREATMENT

Raw water must be free from strong oxidants (such as chlorine) which attack resins. Ideally, water should be free from organics and suspended solids, but practically, ion exchangers can be operated in the presence of both. Strong base resins foul with organics (particularly humic acids) when operating on the OH^- cycle. When operating on the Sul-biSul cycle, organic fouling does not seem to be a problem. Resins are good coagulants and suspended solids cling to them. The problem can be minimized by upflow service through an expanded bed, but this limits the flow rates achievable.

6. EFFLUENT QUALITIES

With the strong acid - weak base system leakage will be 20% or less. The first ions to emerge will be Na^+, HCO_3^- and Cl^-, but if a degasifier is used, the HCO_3^- will be reduced to less than 10 mg/l. At 20% leakage, all of the 500 mg/l TDS water must go through the exchange beds to produce 100 mg/l TDS product.

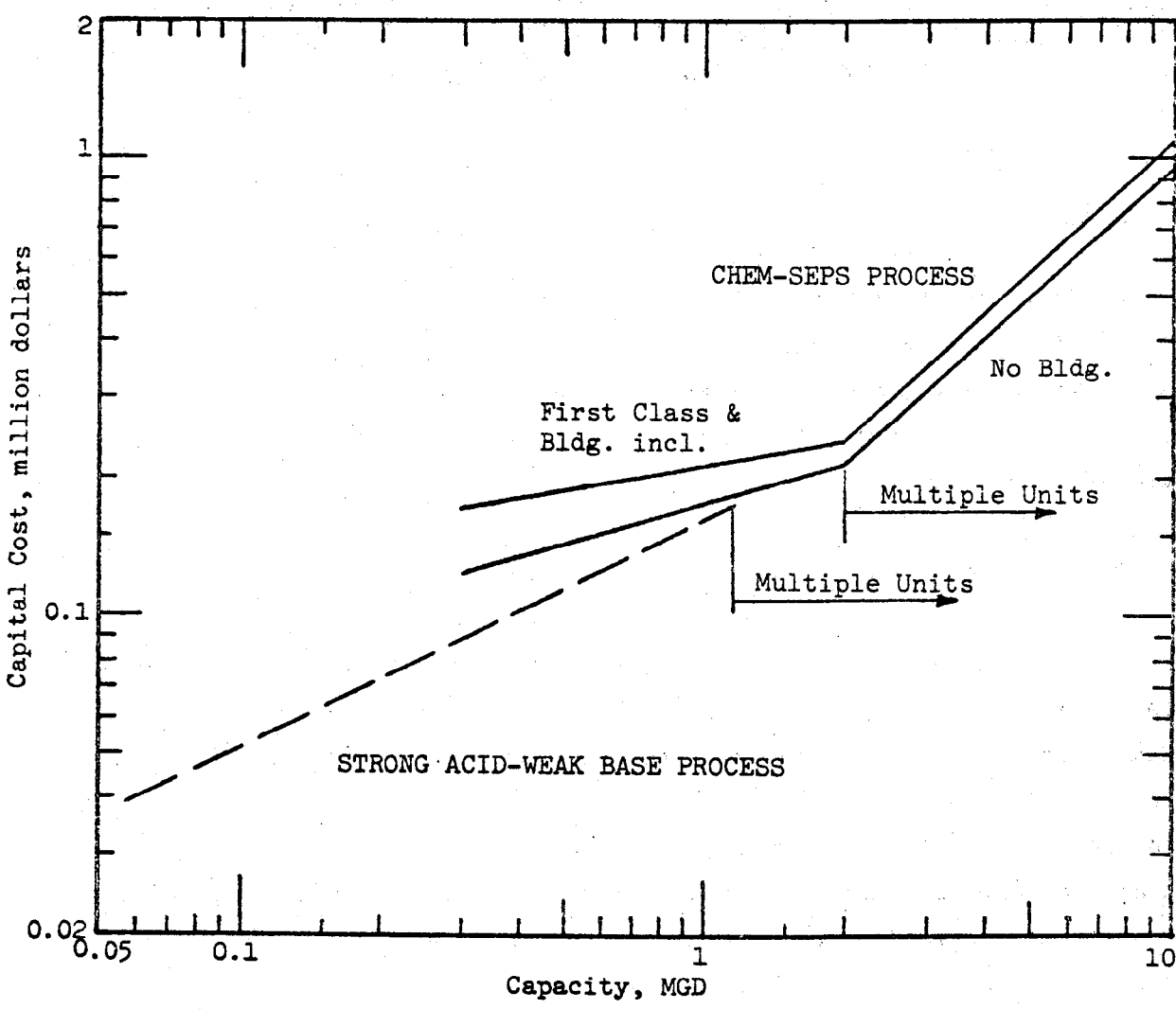


FIG. 8. CAPITAL COSTS FOR ION EXCHANGE PLANTS

The Chem-Seps process can be operated to produce a leakage of about 10 mg/l. Hence, 18% (roughly) can be by-passed and blended to produce a product containing 100 mg/l TDS according to the following calculation.

$$P = FX + L(1-4)$$

$$X = \frac{P-L}{F-L} = \frac{100-10}{500-10} = 0.18$$

where P, F, and L are concentration of TDS in product, feed, and leakage respectively and X is the proportion of by-passed water.

7. EXPERIENCE

In 1965 there were more than 17,000 operational ion exchange plants in the U.S. One (F.E. Weymouth plant) had been softening an average of 302 MGD for a dozen years with almost no loss of resin capacity. Ion exchange is reliable, practical, and there is a vast expertise in design, construction and operation. There are several manufacturers and costs are competitive.

Experience in partial demineralization of high solids (TDS> 500 mg/l) water is much less extensive, but much of ion exchange technology is common to both systems. Thus, even desalination plants are essentially "off-the-shelf" items. However, there is little large scale pilot plant experience with the demineralization of sewage.

A. Strong Acid - Weak Base

Sanks has operated bench scale pilot plants on sewage and high solids water for about five years. There is sufficient evidence to class this system as thoroughly reliable and easy to operate.

B. Sul-biSul Process

A considerable amount of laboratory testing has been done. The only field test is the 0.5 MGD Sul-biSul process operating in a Chem-Seps CCIX

reactor at Burgettstown, Pennsylvania. An acid mine waste is demineralized from 1500 mg/l TDS to 300 mg/l TDS. Cation exchange efficiency is 83%. Anion exchange efficiency is nearly 100% with $Ca(OH)_2$ regenerant. The plant was put into service April 29, 1970. It operates well but the anion exchange requires some readjustment.

C. Desal Process

There has been extensive testing of the Desal process in the Laboratories of Rohm and Haas Co. and in Italy by Sturla (1964). Thibodeaux and Wright (1970) have applied the Desal process to kraft pulp wastes. Kunin (1967), the inventor of the process, has modified it for the treatment of secondary effluents. It was estimated to cost 15¢ to 20¢/1000 gal in a 1 MGD plant which in turn would cost $175,000. Another more extensive modification is being planned to develop a closed system treatment for total recycle of water in a very large ore reduction plant. The minimum economic size of such an installation is 2 MGD. It is estimated that 1800 mg/l TDS can be reduced to 500 mg/l TDS in a 6 MGD plant costing $1,700,000 at an operating cost (exclusive of amortization) of about 16¢/1000 gal. No liquid wastes would be produced.

The Office of Saline Water (1969-1970) reported on the operation of a mobile pilot plant built and operated by The Permutit Company to desalinate waters containing 1000 to 3000 mg/l TDS. The results were successful.

There is little information on the Desal Process applied to sewage. However, Sanks (1970) found IRA-68 resin fouls badly on kraft bleach wastes. Hence, pretreatment by organic carbon may be necessary. As this is a part of both Schemes 1 and 2, there should be no problem with the Desal process.

D. Chem-Seps

There are about 70 Chem-Seps plants operating in the U.S. and another 50 in Europe and Japan. Earlier problems have been solved and the plants now operate with a minimum of attention.

A Chem-Seps pilot plant to treat 10 to 15 gpm of secondary effluent at Pomona, California has just been completed and shipped. No data are yet available.

E. Progressive Mode

A 1000 gpm Progressive Mode softening plant serves Park Service South, Illinois. The Permutit Company has extensively tested an excellent, fully automated desalination pilot plant with good results. There is reason to think a large-scale plant would perform as well.

F. Selective Ion Exchange

A mobile pilot plant to remove ammonia from secondary sewage treatment plant effluents was built and operated by Battelle Memorial Institute (1969). The plant contained facilities for flocculation, sedimentation, carbon sorption, disinfection, mixed media filtration, and ion exchange using Hector clinoptilolite. Ammonia removal of 97% from wastewater was achieved at a flow rate of 70,000 gpd through three 750 gallon ion exchange reactors. The results showed this method to be highly effective.

8. CAPITAL AND OPERATING COSTS

Approximate costs are given for a Chem-Seps plant to reduce 500 mg/l TDS to 100 mg/l in a 10 MGD plant.

Capital cost (Item 9B) = $950,000	
Operating Cost	
Amortization (Item 9C)	2.5¢/1000 gal
Acid regenerant (Item 9D)	5.1
Base regenerant (Item 9E)	1.9
Resin replacement	
Cation, 10%/year + anion, 20%/year	1.0
Maintenance	0.8
Power	0.7
Labor, total	1.4
Total cost, ¢/1000 gals	13.4¢/1000 gal

These costs are low compared to costs estimated for other processes by the Pacific Southwest Inter-agency Committee (1970) but nevertheless they are thought to be correct to within 3¢ to 6¢/1000 gal. Very large quantities of acid and lime can probably be obtained at less cost than those used herein. Labor may cost more. But in conclusion, the cost given is within reason.

9. SAMPLE CALCULATIONS

A. IX Capacity: by-pass 18% water, so provide for 8.2 MGD plus 3% waste = 8.5 MGD

B. See Fig. 8. Total plant cost = $950,000

C. Amortize at 7% over 20 years

$$R = P\left[\frac{i(1+i)^n}{(1+i)^n-1}\right]$$

$$R = 950{,}000\left[\frac{0.07(1.07)^{20}}{(1.07)^{20}-1}\right] = \$89{,}600/\text{yr}$$

$$R = \$89{,}600 \times \frac{1}{365} \times \frac{1{,}000\ \text{gpd}}{10{,}000{,}000\ \text{gpd}} = 2.5¢/1000\ \text{gal}$$

D. Acid regenerant

NB: Removal of 400 mg/l TDS = 6.1 me/l cations (See Table III)

Acid efficiency = 90%

Acid use = $\frac{6.1}{0.9}$ = 6.8 me/l = 0.0068 eq/l

Per 1000 gal:

$$0.0068\ \frac{\text{eq}}{1} \times \frac{3.79\ \text{l}}{\text{gal}} \times 1000\ \text{gal} \times \frac{98\ \text{g}}{2\ \text{eq}} \times \frac{\text{lb}}{454\ \text{g}} \times \frac{1\ \text{ton}}{2000\ \text{lb}} = 0.0014\ \text{tons}$$

H_2SO_4 = 0.0015 tons 93.6% acid

Cost @ $34/ton = $34 x .0015 = $0.051/1000 gal.

E. Base regenerant

NB: Removal of 400 mg/l TDS = 6.1 me/l anions, but 30% is bicarbonate removable by degasification of cation effluent leaving 6.1 x 70% = 4.3 me/l strong electrolyte anions to be removed

Base efficiency = 90%

Base use = $\frac{4.3}{0.9}$ = 4.8 me/l = 0.0048 eq/l

Per 1000 gal:

$$0.0048 \frac{eq}{l} \times \frac{3.79\ l}{gal} \times 1000\ gal \times \frac{74\ g}{2\ eq} \times \frac{lb}{454\ g} \times \frac{1\ ton}{2000\ lb}$$

$$= 0.00074 \text{ tons } Ca(OH)_2 = 0.00077 \text{ tons of } 96\% \ Ca(OH)_2$$

Cost @ \$24/ton = 24 x .00077 = \$0.0185/1000 gal

10. SCALE-UP FACTORS

Beyond the maximum practical reactor size, multiple units must be used, and they are actually desirable for operational flexibility. The cost of n units is n times the cost of a single unit with a discount of 15% for duplication.

In operational characteristics, there are some differences between bench-scale pilot plants and prototypes. Flow patterns, tailing, and gravitational currents are different in large reactors and lead to slightly less efficient operation in the prototype.

11. PROCESS DEPENDABILITY

Ion exchange is a highly dependable process but it can be upset by strong oxidants (which attack the resins), by coatings of particles or slime (which blind the resin), or by irreversible exchange (which "poisons" the resins). But ion exchangers have been operated on contaminated waters, and when the ion exchangers are protected by flocculation and carbon sorption, difficulties have been minimal.

12. MATRIX SUMMARY

Ion exchange is a mature process. While there may be changes and improvements, there seems to be little reason to expect revolutionary breakthroughs. In the past two decades, improvements and competition have kept the cost from rising as rapidly as the Engineering News Record Construction Cost Index. One might speculate this will continue. One

might also speculate that sulfuric acid (a major item of cost) will decrease in cost when scrubbers to remove SO_2 from stack gases from soft coal combustion become common.

With respect to the items in the matrix, there is little or no difference anticipated in the years 1975, 1990, and 2020.

Scale-Up	Straight line above 2-4 MGD
Dependability	Good
Flexibility	Excellent
Tolerance to feed	Good except oil and grease and oxidants must be trace quantities or less. SS must be reasonably low.
Effluent variability	Low
Ancillary processes required	Coagulation and sorption
Costs	Will rise somewhat less than the pace of inflation
Efficiency	85% to 95%
Operational Difficulty	Requires expert for start-up. Control maintenance requires expert. Operation is automatic and, once established, is easy.

LIST OF REFERENCES

Anderson, J. A. and N. A. Voss. "Progressive-mode continuous countercurrent ion exchange." Industrial Water Engineering (June 1970).

Battelle Memorial Institute. "Ammonia removal from agricultural runoff and secondary effluents by selected ion exchange." Robert A. Taft Research Center Report No. TWRC-5. U.S. Dept. of the Interior (March 1969).

Kunin, R. "Applications of ion exchange." Amber-Hi-Lites 100 (Philadelphia, Rohm and Haas Co., July 1967).

Kunin, R., "Applications of ion exchange." Amber-Hi-Lites 116. (Philadelphia, Rohm and Haas Co., March 1970).

Neale, J. H. "Advanced waste treatment by distillation," AWTR-7, U.S. Public Health Service Dept. 999-WP-10, Robert A. Taft Sanitary Engineering Center, Cincinnati, Ohio (1964).

Office of Saline Water. "1969-1970 Saline Water Conversion Report." U.S. Dept. of the Interior, (1969-1970).

Pacific Southwest Inter-agency Committee. "An inventory of wastewater reclamation and re-use in the Pacific Southwest area." (1970).

Sanks, R. L. and W. J. Kaufman. "Recycling of saline wastes," Advances in Water Pollution Research, 3:333-359, (c 1967).

Sanks, R. L. "Improving the efficiency of cation exchange," Proc. American Power Conference, 29:808-814, (1967).

Sanks, R. L. "Color and mineral removal from kraft bleach wastes, Progress Report 4" (unpublished) to Water Quality Office, Environmental Protection Agency Project 12040 DBD, (November 1970).

Sturla, P. "Pilot plant studies of the Kunin process," presented at Int. Water Conf. of the Engineer's Soc. of Western Penna., Pittsburgh, Penna., (September 1964).

Thibodeaux, L. J. and R. L. Wright. "Deionization of paper industry waste water via ion exchange." Proposed for a Technical Bulletin of the National Council of the Paper Industry for Air and Stream Improvement Inc. In press.

Valentine, Dennis. Corps of Engineers. Private Communication, (May 6, 1971).

REPORT ON ADSORPTION FOR THE MERRIMACK BASIN PILOT WASTEWATER MANAGEMENT PROGRAM

W. J. Weber, Jr.

College of Engineering

The University of Michigan

Ann Arbor, Michigan

May 1971

epared for the U.S. Army Corps of Engineering, North Atlantic Division

REPORT ON ADSORPTION FOR THE MERRIMACK BASIN PILOT WASTEWATER MANAGEMENT PROGRAM

1. Definition of the Problem

This report on adsorption technology is part of an overall "maximum water quality" treatment feasibility program being conducted by the U.S. Army Corps of Engineers on the Merrimack River Basin of Massachusetts and New Hampshire. The program is part of a large pilot wastewater management program, with other similar activities being conducted in Chicago, Cleveland, Detroit, and San Francisco.

Three different regions of widely different character have been selected for the Merrimack Basin Pilot Program (MBPP); namely; the Boston Area, the Lowell-Lawrence-Haverhill area, and the Winnepesaukeee area. The regions vary from rural to urban and include some of the more difficult and concentrated industrial wastes to treat. These industries include the classical water quality problems associated with tanneries, textile, metal plating, and paper manufacturing as well as the more recent discharges of plastic wastes. The wastewater collection systems range from old combined sewers to recently constructed separate systems. The domestic waste component varies from relatively small volumes in some areas of the basin to significant volumes in the major urban complexes. The objective is to design a feasible treatment system. This particular report is addressed to the question of how adsorption, as an advanced unit peration for waste treatment, can be best used in the "most feasible" system to be composed."

This report represents a preliminary contribution to an overall treatment program feasibility report which will provide the basis for discussion at a mid-May 1971 Workshop in Amherst, Massachusetts. The format of the report follows directly an outline suggested by the Corps.

2. Concept and Theory - Summary

Historically, the occurrence of adsorption was first noted by C. W. Scheele in 1777; Scheele observed the selective removal of gases from air by charcoal. Adsorption of coloring matter from aqueous solution by charcoal was later observed by T. Lowitz in 1791. Today it is recognized that adsorptive reactions are prevalent in most natural physical, biological and chemical processes, and adsorption on solids such as active carbon has become widely used operation for purification of waters and wastewaters.

Adsorption is a process involving the inter-phase accumulation or concentration of substances at a surface or interface. Adsorption can occur at an interface between any two phases, for example a liquid-liquid interface, a gas-liquid interface, or a liquid-solid interface. The material being concentrated or adsorbed is referred to as the adsorbate, and the adsorbing phase is called the adsorbent.

Absorption, vis-a-vis adsorption, is a process in which the molecules or atoms of one phase penetrate nearly uniformly among those of another phase to form a solution with the second phase. The term sorption, which includes adsorption and absorption, is thus a very general expression for a process in which a component moves from one phase to another, particularly for cases in which the second phase is solid. The phenomenological differences between adsorption and absorption, and chemical compound formation are illustrated graphically in Figure 1 for reactions of each type in which a substance moves from liquid phase to a solid phase.

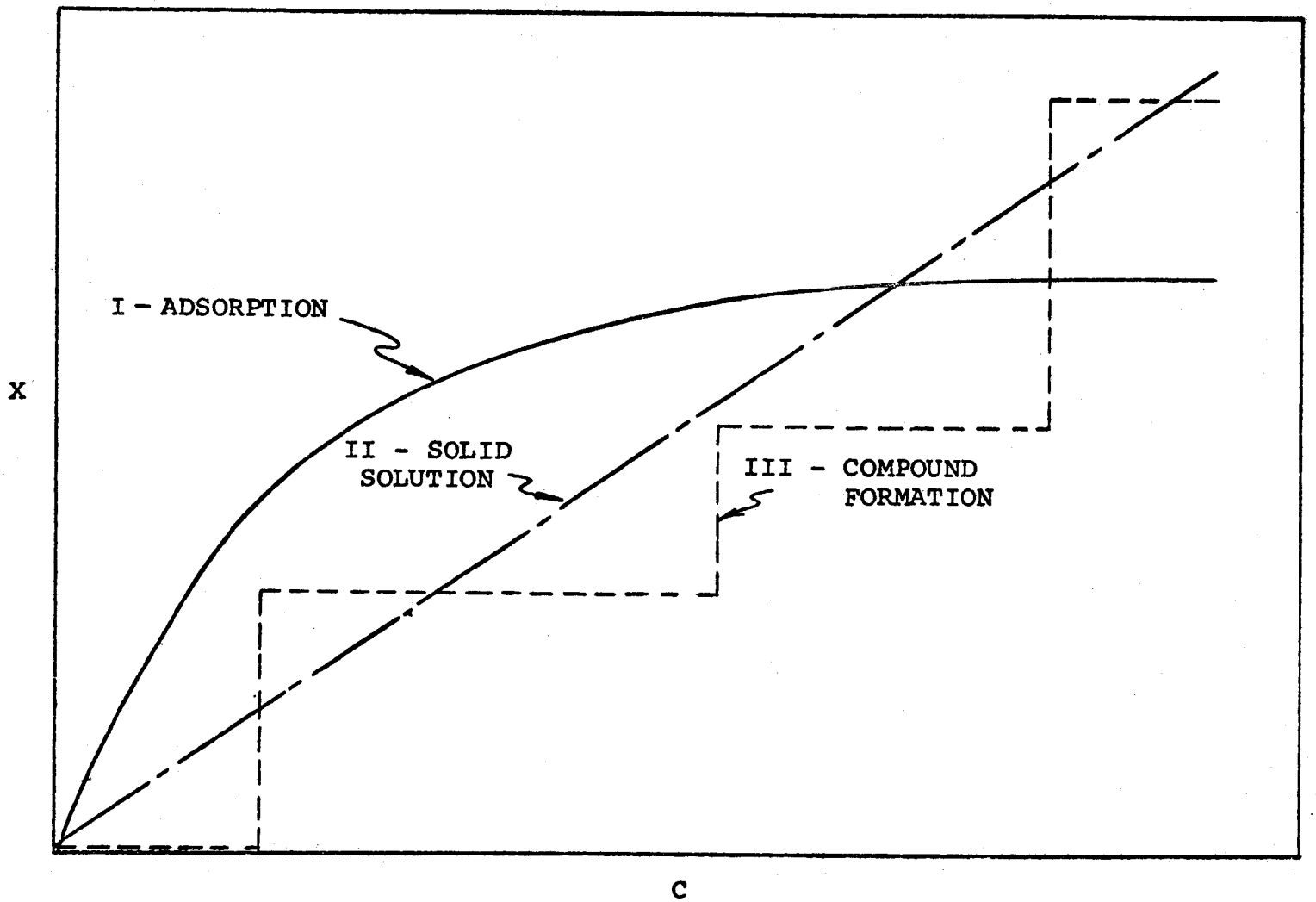


Figure 1. Types of Solid State Interactions.

The term, C, on the abscissa of Figure 1 represents the concentration of the migrating substance in the liquid phase, which is in contact with the solid phase to form a liquid-solid interface. A granule of activated carbon suspended in a solution of phenol can serve as an example of this type of system. The term, X, on the ordinate then denotes that amount of the substance that has moved across the interface. Curve I indicates, for adsorption, the curvilinear dependence of the amount concentrated at the solid surface on the amount in the solution phase. Curve II represents the absorption of molecules of the substance from solution by the solid phase to form a solid solution in direct proportion to the concentration of the substance in the liquid phase. Finally, Curve III demonstrates the step-function relationship between the reacted material in the solid phase and the unreacted material in the liquid phase for the process of chemical compound formation.

Causes and Types of Adsorption

The process of adsorption from solution on a solid occurs as the result of one of two characteristic properties for a given solvent-solute-solid system, or a combination thereof. The primary driving force for adsorption may be a consequence of lyophobic (solvent-disliking) behavior on the part of the solute relative to the particular solvent in question, or of a high affinity of the solute for the solid. For the majority of systems encountered in water and waste-water treatment practice, adsorption results from the combined action of the two forces.

Positive adsorption in a solid-liquid system results in the removal of solutes from solution and their concentration at the surface of the solid. This process proceeds in a given system to such time as the concentration of the solute remaining in solution is in a dynamic balance with that at the surface. At this position of dynamic balance, or equilibrium position, there exists a defined distribution of solute between the liquid and solid phases. The distribution ratio is a measure of the position of equilibrium in the adsorption process; it may be a function of the concentration of the solute, the concentration and nature of competing solutes, the nature of the solution, etc. The preferred form for depicting this distribution is to express the quantity X as a function of C at fixed temperature, the quantity X being the amount of solute adsorbed per unit weight of solid adsorbent, and C the concentration of solute remaining in solution at equilibrium. An expression of this type is termed an <u>adsorption isotherm</u>. Basically, the adsorption isotherm is a functional expression for the variation of specific adsorption with concentration of adsorbate in bulk solution at constant temperature. Commonly, the amount of adsorbed material per unit weight of adsorbent increases with increasing concentration, but not in direct proportion (see Fig. 1, curve I).

The adsorption isotherm is not only the most convenient form for representing experimental data at a given temperature, but is also a useful starting-point for the development of theoretical

treatments of adsorption equilibria. One of the primary conditions that an adsorption theory must fulfill is that it adequately describe the experimental isotherm. Satisfaction of this condition, however, does not in itself constitute validation of an adsorption theory.

Several types of isothermal adsorption relations may occur. The most common relationship between X and C obtains for systems in which it appears that adsorption from solution leads to the deposition of only a single layer of solute molecules on the surface of the solid. Occasionally, however, multimolecular layers of solute may be adsorbed. Then, for adequate description of the phenomenon, resort must be had to adsorption models which are somewhat more complex.

The Langmuir and B.E.T. Equations

The familiar Langmuir model is valid only for single-layer adsorption, while the Brunauer, Emmett, Teller - B.E.T. - model will represent many isotherms reflecting apparent multilayer adsorption. Both equations are limited by the assumption of uniform energies of adsorption on the surface. Graphically the Langmuir-model isotherm has the form shown in Figure 2a, and the most common B.E.T. - model isotherm is represented by Figure 2b; in both drawings the saturation concentration of the solute in the solution at a given temperature is represented by C_s. The B.E.T. isotherm, the more generally applicable of the two, reduces to the Langmuir model when the limit of adsorption is a monolayer.

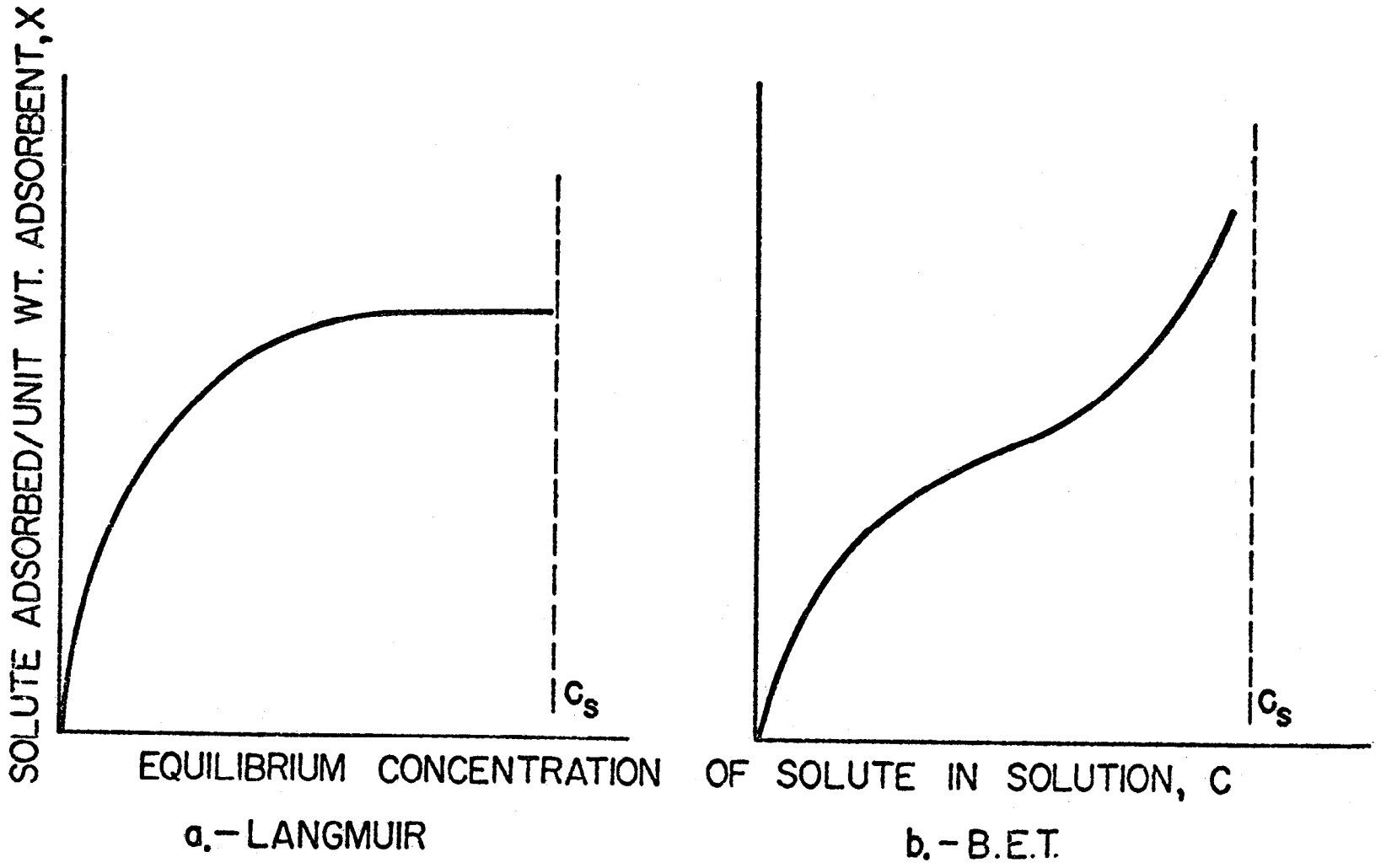


Figure 2. Typical Isotherms for Langmuir and B.E.T. Adsorption Patterns.

Both the Langmuir and the B.E.T. isotherms may be deduced from either kinetic considerations or the thermodynamics of adsorption. The latter derivations are somewhat more sophisticated, though less intuitive, than the kinetic treatments since fewer assumptions are involved (e.g., the balancing of forward and reverse rate processes according to some assumed mechanism). Derivations of these isotherms are readily available in the literature. The Langmuir treatment is based on the assumptions that maximum adsorption corresponds to a saturated monolayer of solute molecules on the adsorbent surface, that the energy of adsorption is constant, and that there is no transmigratory activity of adsorbate molecules in the plane of the surface. The B.E.T. model assumes that a number of layers of adsorbate molecules form at the surface and that the Langmuir equation applies to each layer. A further assumption of the B.E.T. model is that a given layer need not complete formation prior to initiation of subsequent layers, hence the equilibrium condition will involve several types of surfaces in the sense of number of layers of molecules on each surface site at its steady state value.

For adsorption from solution with the additional assumption that layers beyond the first have equal energies of adsorption, the B.E.T. equation takes the simplified form

$$X = \frac{ACX_m}{(C_s - C)\ \left(1 + (A-1)\frac{C}{C_s}\right)} \qquad (2)$$

in which C_s is the saturation concentration of the solute, C is the measured concentration in solution at equilibrium, X_m is the

number of moles of solute adsorbed in forming a complete monolayer on the carbon surface, X(or x/m) is the number of moles of solute adsorbed per gram of carbon at concentration C and A is a constant expressive of the energy of interaction with the surface. Equation (2) may be rearranged to facilitate its application to experimental data, giving the linear form

$$\frac{C}{(C_s-C)X} = \frac{1}{AX_m} + \frac{A-1}{AX_m} \cdot \frac{C}{C_s} \qquad (3)$$

A plot of the left hand term of Equation (3) against C/C_s gives a straight line of slope $(A-1)/AX_m$ and intercept $1/AX_m$ for data which accord with the B.E.T. model. A plot of this type is illustrated in Figure 3a.

The Langmuir isotherm is

$$X = \frac{X_m bC}{(1+bC)} \qquad (4)$$

in which b is a constant related to the energy of adsorption and all other symbols have the same significance as in Equations (2) and (3). Equation (2) will reduce to Equation (4) if b is set equal to A/C_s, C is taken as negligibly small compared with C_s, and A is taken as much greater than 1 . Two convenient linear forms of the Langmuir equation are

$$\frac{C}{X} = \frac{1}{bX_m} + \frac{C}{X_m} \qquad (5)$$

or

$$\frac{1}{X} = \frac{1}{X_m} + \frac{1}{bX_m} \cdot \frac{1}{C} \qquad (6)$$

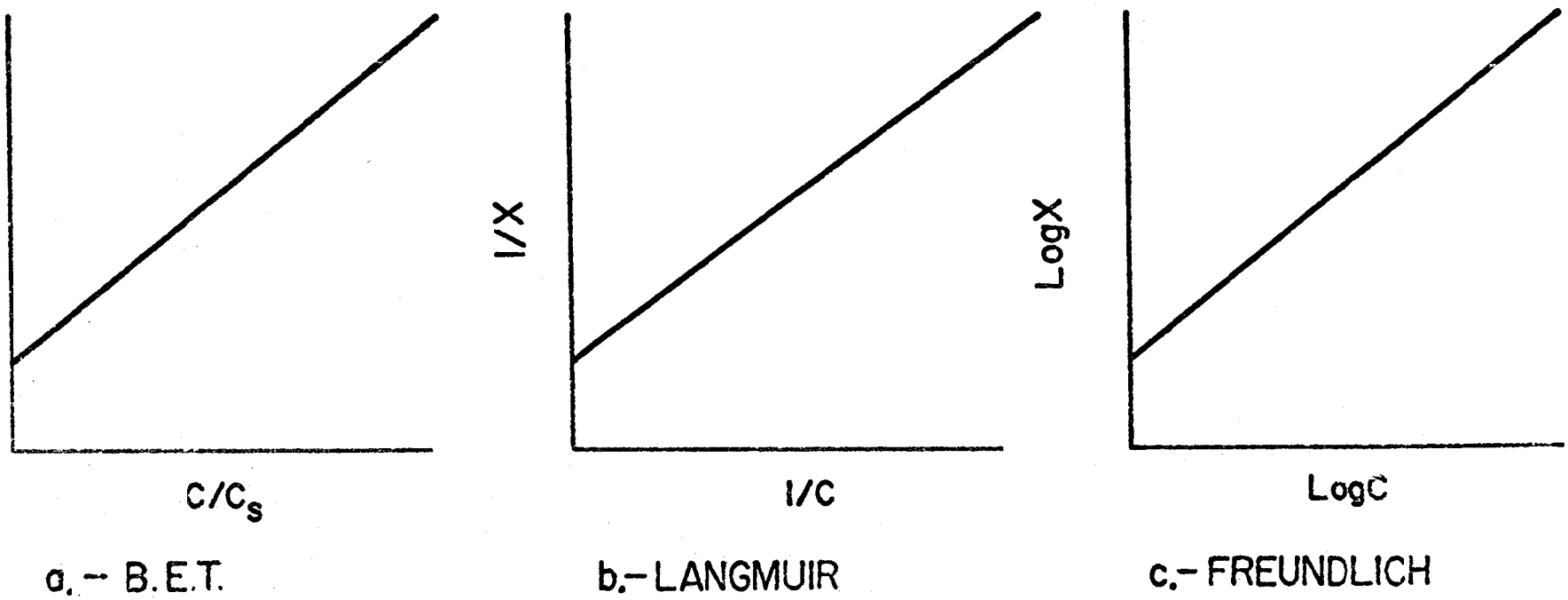


Figure 3. Linear Forms for Graphical Representation of Adsorption Equations.

Either of these forms may be used for linearization of data that accord with the Langmuir equation. The form chosen usually depends on the range and spread of the data and on the particular data to be emphasized. The type of plot used for Equation (6) is illustrated in Figure 3b.

For very small amounts of adsorption, that is, when $bC \ll 1$, the specific adsorption is proportional to the final concentration of adsorbate in solution, then

$$X = X_m bC \tag{7}$$

For large amounts of adsorption

$$bC \gg 1, \quad \text{and}$$

$$X \approx X_m \tag{8}$$

When adsorption is in accord with the Langmuir equation the total capacity of the adsorbent for an adsorbate is given by the limiting value of X as C approaches C_s and is equal to the value of X_m. Assumption of a value for the surface area covered per molecule then allows computation of the active specific surface area of the adsorbent. The Specific surface area, Σ_s, can be determined by the relationship

$$\Sigma_s = X_m N \sigma^o \tag{9}$$

in which σ^o is the area per molecule, N is Avogadro's number, and X_m is expressed in moles of adsorbate per unit weight of adsorbent.

The Freundlich Equation

One other equation for isothermal adsorption, the Freundlich or van Bemmelen equation, has been widely used for many years.

This equation does not derive from theoretical concepts, as do the Langmuir and B.E.T. equations, but is nonetheless quite useful. The Freundlich equation, which has the form

$$X = KC^{1/n} \tag{10}$$

is empirical and contains two adjustable constants, K and n. Data are usually fitted to the logarithmic form of the equation

$$\log X = \log K + \frac{1}{n} \log C \tag{11}$$

which, as illustrated in Figure 3c, gives a straight line with a slope of $1/n$ and an intercept equal to the value of $\log K$ for $C = 1$ ($\log C = 0$).

The Freundlich equation is commonly used for comparison of powdered carbons for water-treatment usage. The amount adsorbed, X, is equal to $(C_i - C)/Vm$, C_i being the initial concentration of phenol or other contaminant and V the volume of water treated. The value for m is then given by D/V, D being the dosage of carbon, i.e., weight per unit volume. Substitution into Equation (11) gives

$$\log \frac{C_i - C}{D} = \log K + \frac{1}{n} \log C \tag{12}$$

3. Applicability ot the Systems in Question

Applications of adsorption on active carbon for purification of public water supplies have been well known and widely practiced for years. The taste and odor producing impurities removed in such applications are usually present in low concentrations, and the need to provide removal often intermittent or occasional. Consequently, a commonly suitable mode of application is one in which an appropriate quantity of carbon is introduced to a given volume of water, held in contact with the water for a specified period, then separated and discarded. The form of carbon used for such operations is usually powdered, largely for reasons of lower initial cost and adaptability to intermittent operation.

It has been clear from inception of the Advanced Waste Treatment Research Program in the Division of Water Supply and Pollution Control of the Public Health Service in the early 1960's that the role of adsorption as a unit process for providing high levels of treatment for wastewaters would be significant, for purposes of both water pollution control and for reclaiming wastewaters for reuse (Morris and Weber, 1962; PHS, 1962 and 1964). It has been just as clear that the relatively inefficient operations commonly used for water purification are not satisfactory for waste treatment applications. With relatively large amounts of organic impurities to remove, and with the need for continuous treatment, much more efficient utilization of the capacity of active carbon is required. Further, the fact that large

amounts of carbon are needed for such operations requires a scheme of regeneration and reuse of the carbon. These requirements strongly suggest the use of continuous contacting systems.

The most common type of continuous contacting unit is one in which a stream of wastewater is passed through a bed of granular carbon. The granular form is suggested by the relative ease of handling and regenerating this material relative to powdered carbon. In a bed or columnar contacting system the carbon in any section of the bed is continuously in contact with a solution of increasing concentration of impurities. Thus, there is no double-effect leveling of the driving force for adsorption. Further, if the bed is fixed, the maximum advantage of the gradient driving force is realized, in that the most saturated carbon is in contact with a solution having the highest concentration of impurities while the least saturated carbon is in contact with a solution having the lowest concentration of impurities.

The potential of adsorption on granular carbon for "tertiary" treatment of wastewaters has been demonstrated, (Johnson et al, 1964; Joyce and Sukenik, 1964 and 1965; Weber and Atkins, 1966; Weber, 1967) and there are several large scale pilot projects underway. Notable among these are the studies of packed-bed contacting systems at Lake Tahoe (Schlecta and Culp, 1967) and Pomona (Parkhurst et al, 1967), California and Nassau

County, New York, where adsorption on granular active carbon is one of a series of processes used to treat secondary effluent to produce water for recharge of ground water aquifers (Stevens and Peters, 1966). The use of powdered carbon for renovation of secondary effluent has been studied by Davies and Kaplan (1964) and Beebe and Stevens (1967), among others.

Packed beds of granular active carbon are well suited for treatment of solutions that contain little or no suspended solids, and under such circumstances can normally be expected to operate effectively for extended periods without clogging or excessive pressure loss. However, the suspended solids invariably present in municipal and industrial wastewaters present some problems for the use of packed beds. These solids lead to progressive clogging of the carbon beds, much as they do in sand filters, with resulting increases in head loss. At the Pomona installation the filtering action of packed beds of active carbon is used as part of the treatment scheme for removal of suspended solids. In this sytem the first of four active carbon beds operating in series serves as a filter, providing clear feed to the three subsequent beds. Suspended solids are removed from the first bed daily by surface washes and backwashing. The Lake Tahoe and Nassau County systems include pretreatment of secondary effluent by chemical clarification and filtration, providing a highly clarified feed and permitting extended operation of the carbon beds (Schlecta and Culp, 1967; Stevens and Peters,

There is an increasing emphasis on the potential of expanded-bed adsorbers, which have certain inherent operating advantages over packed-bed adsorbers for treating solutions which contain suspended solids. By passing wastewater upward through a bed of activated carbon at velocity sufficient to expand the bed, problems of fouling, plugging and increasing pressure drop are eliminated. Effective operation over longer periods of time results, as has been demonstrated clearly in comparative laboratory studies and preliminary small-scale field studies of expanded-bed adsorbers (Weber, 1967) and in larger-scale pilot field investigations in both "tertiary" (Weber et al, 1968) and direct physicochemical (Weber et al, 1970) applications. Another advantage of the expanded bed is the lack of dependence of pressure drop on particle size. It is possible to use carbon of smaller particle size in an expanded bed than is practical in a packed bed, thus taking advantage of the high adsorption rates which obtain for smaller particles (Weber and Morris, 1963).

It has become apparent over the past several years that achievement of high levels of water quality demanded by progressive water use and reuse requirements, and by requirements for more effective water pollution control, will necessitate expanded utilization of advanced technologies for wastewater treatment. Conventional "secondary" biological treatment processes do not provide the degree of treatment required for most water reuse applications, nor will they provide a completely satisfactory long-term measure for protecting natural waters from pollution by waste discharges.

Well operated modern biological waste treatment plants can provide approximately 90 % removal of suspended solids and biochemical

oxygen demand (BOD). Although the quality of the effluent from such plants has been adequate to meet most discharge regulations and standards in the past, increases in population and in standard of living have resulted, in the face of a relatively fixed total water resource, in more stringent demands for better water quality and more effective pollution control. As a result, significant interest has focused over the past decade or so on development of physicochemical processes capable of accomplishing the degree of treatment required by more stringent effluent standards.

Common philosophy regarding application of advanced physicochemical processes for wastewater treatment initially centered on providing "tertiary" treatment for wastes which had undergone conventional "secondary" biological treatment. The addition of tertiary-level physicochemical processes to conventional biological processes incurs significant additional treatment expenses. Further, the effective operation of a teritary treatment system depends on consistent and efficient operation of the biological secondary process, which remains subject to problems arising from changes in waste composition, from large variations in flow which often have to be diverted, and from the presence of toxic materials which disrupt biological oxidation processes.

Most recently, physicochemical treatment has been developed and successfully demonstrated as an attractive technical and economic alternative to biological treatment for direct application to raw wastewaters (Weber et al, 1970; Hager and Reilly, 1970). For this application the basic physicochemical treatment process consists of adsorption by activated carbon preceded by chemical clarification,

and possibly accompanied by filtration.

The concept of applying the process directly to a primary waste rather than to a secondary effluent derives partially from observations regarding the apparent difficulty of removing final traces of organic material from secondary effluents by treatment with activated carbon, as well as from the relative economics of two-stage vis-a-vis three-stage treatment systems.

Consistent organic matter removals of 95 to 97 percent have been maintained, despite variations in waste strength and composition. Toxic substances which adversely affect biological treatment processes have little or no effect on the physicochemical process. The effluent produced by the physicochemical process is essentially free of suspended solids and contains only about 5 mg/l or less of total orgnaic carbon (TOC) and biochemical oxygen demand (BOD), compared with average TOC and BOD values of about 30 to 35 mg/l for the same wastewaters treated conventionally.

In addition, the physicochemical treatment process generally achieves a high degree of removal of phosphate (about 90%) and nitrate, inorganic algal nutrients which normally are not removed effectively by conventional biological wastewater treatment.

4. Design Criteria

The primary objective of waste treatment is sufficient reduction in the level of pollutants in the wastewater to allow safe discharge to the environment. The pollutants in sewage commonly are grouped into classes of similar compounds which have the same environmental impact. Presently the five major pollutant groups of interest are: suspended solids, organic matter as measured by BOD, TOC, or COD, phosphorus compounds, nitrogen compounds and pathogenic organisms. It is not possible to stipulate one set of effluent standards which is applicable to all situations. However, it is generally agreed that a good quality effluent will have the characteristics given in Table 1. In addition, complete disinfection is normally required. There is not yet any general agreement on a nitrogen level. The quality parameters listed in Table 1 are not design standards, but are guides for describing a good quality effluent.

Table 1 - Desirable Effluent Quality

BOD	$\leq$	10 mg/l
COD	$\leq$	30 mg/l
Suspended Solids	$\leq$	10 mg/l
Phosphorous	$\leq$	1 mg/l

The role of carbon adsorption is removal of soluble organics from wastewater. Although preceding treatment by chemical clarification does the bulk of the pollution control job, the carbon adsorption step is required to produce a good quality effluent. The results obtained by the combination of clarification and carbon adsorption at several physical-chemical pilot plants have illustrated that not only is the

removal achieved quite high (95%+) but equally important the residual organics after treatment are quite low. These effluents are superior to the usual quality of secondary effluent (TOC ≈ 20 mg/l, COD ≈ 40 to 50 mg/l).

Best results to date have been obtained with carbon contacting systems employing granular carbon. In such systems the waste water is passed either upward or downward through columns containing the granular carbon. Downflow columns function additionally as packed beds to accomplish some filtration of the wastewater. Flow rates of 2 gpm/ft^2 to 8 gpm/ft^2 have been employed. In this flow range essentially equivalent adsorption efficiency is obtained provided the same contact time is employed. At flow rates below 2 gpm/ft^2 adsorption efficiency is reduced, while at flow rates above 8gpm/ft^2 excessive pressure drop takes place in packed beds. Contact times employed are in the range of 30 minutes to 60 minutes on an empty bed basis. In general, increases in contact time up to 30 minutes yield proportionate increases in organic removal. Beyond 30 minutes the rate of increase falls off with increases in contact time and at about 60 minutes contact time becomes negligible. Carbon beds to be operated at the lower end of the flow rate range are generally designed for gravity flow. Those systems designed for the higher flow rates must employ pressure vessels if packed beds are used. A pressure vessel is more expensive to construct than a gravity flow vessel but it requires less land area, and provides greater ability to handle fluctuations in flow rate.

Provision must be made to regularly backwash packed bed carbon beds because they collect suspended solids. In addition, biological

growth tends to clog such beds. It is advisable to include a surface wash and air scour to be assured of removal of gelatinous biological growth.

Backwashing satisfactorily relieves clogging but does not completely remove biological growth. Consequently, significant biological activity is manifest. This leads to the development of anaerobic conditions in the packed carbon beds and generation of sulfides. Aeration of the column feed to packed beds has been utilized to prevent anaerobic conditions, but his produces so much biological growth that excessive backwash is required.

To overcome these difficulties upflow expanded bed carbon columns have attracted increased interest. These beds allow for significant accumulation of biological activity with little increase in head loss. Consequently, aerobic conditions can be maintained and sulfide generation prevented.

The exhaustion capacity of an active carbon is a critical aspect of the design of any carbon contacting system. For industrial applications this is evaluated by running adsorption isotherm tests. In the wastewater treatment field isotherms are of limited utility because of biological activity which develops on the carbon and tends to enhance its apparent capacity for organic removal.

The design engineer must develop a system which makes greatest utility of the available capacity of the carbon. To provide a good effluent and utilize most of the available capacity, countercurrent contact is commonly required. This can be achieved by having the waste flow through a number of contactors or stages in series in one direction, while the carbon moves in the opposite direction. In powdered

carbon contacting systems this is exactly the procedure used. With granular carbon the procedure cannot be used per se, an undesirable attrition losses will take place. Rather, when exhausted, the lead contactor in a series of adsorption columns is removed from service and a spare contactor with fresh carbon placed at the end of the line. Each contactor is then moved up one position in the line. This is accomplished by piping and valving a series of columns to shift the inflow and outflow points of the series accordingly rather than physically moving the columns. As the number of stages increases, the piping and valving arrangement becomes more complex and costly. A compromise between the advantage of adding another stage to more closely approach the highest use of the carbon capacity, and the cost of each additional stage must be achieved.

Table 2 reviews the various factors which must be considered in designing any carbon removal system.

Table 2

Considerations in Carbon Treatment Design

1. Type of Carbon-Granular or Powered
2. Contact Time
3. Flow Rate
4. Configuration-Series or Parallel
5. Number of Stages
6. Flow Direction - Packed or Expanded
7. Hydraulic Force - Pumped or Gravity
8. Organic Capacity

Table 3 gives carbon capacities obtained in field operation at several physicochemical pilot plants. In view of the fact that the waste, effluent criteria, number of contact stages etc., varied from plant to plant it is not surprising that some spread in the results is observed. These capacities are expressed as pounds of organics removed (either as COD or TOC) per pound of carbon. For general planning purposes a capacity of 0.5 pounds of COD per pound of carbon is reasonable. This is approximately equivalent to a requirement of 500 pounds of activated carbon per million gallons of sewage treated.

Table 3

Carbon Capacity in Physicochemical Treatment Plants

Plant	Carbon Capacity	
	lbs TOC / lb A.C.	lbs COD / lb A.C.
Blue Plains (Wash.)	0.15	0.41
Ewing-Lawrence (New Jersey)	0.3	(0.75 - 0.90)
New Rochelle (New York)	(0.20-0.24)	0.6
Lebanon (Ohio)	0.22	0.5

This carbon requirement would be prohibitively expensive if regeneration and reuse of the carbon were not possible. A technically and economically feasible method is available for regeneration of granular activated carbon by heating in a multi-hearth furnace in the presence of steam to ≈ 1750°F. This treatment burns away adsorbed organics. During each regeneration cycle some carbon is lost by burning and attrition, and some by alteration of surface properties. The overall loss expressed as percent by weight of virgin carbon required to restore the total original capacity of the batch ranges

from 5 to 10 percent. For planning purposes, carbon make-up requirements range from 25 to 50 lbs per million gallons of wastewater can be considered to treated.

At present, experiments on a fluidized bed regeneration system for powdered carbon are moving into the large-scale pilot stage. Successful completion of these tests will be a big step forward in making a powdered carbon system a technical and economic reality. The key factor will be maintaining the carbon loss at a low enough level during the regeneration.

5. Pretreatments Required

Adsorption systems require pretreatment by chemical clarification as discussed previously. For highest water quality and best operating effectiveness, packed-bed adsorbers should be preceded by filtration, while expanded bed adsorbers should be followed by filtration.

6. Effluent Qualities

This subject is covered in a combined report by the project coordinator.

7. Experience on Process to Date

Several prominent examples of the use of adsorption in a "tertiary treatment" capacity have been given in section 4 of this report. This section will thus focus on experiences with adsorption in direct "physicochemical" treatment systems.

Table 4 gives overall results obtained at four different pilot plant installations of physicochemical treatment.

Table 4

Physicochemical Treatment Plants

Plant	Organic Removal %	Effluent Concentration
Ewing-Lawrence (N.J.)	95-98	TOC = 3-5
Blue Plains (Wash., D.C.)	95-98	TOC = 6
Lebanon (Ohio) (powdered carbon)	95	TOC = 11
New Rochelle (N.Y.)	95	COD = 8
Rocky River (Ohio)	93	BOD = 8
Salt Lake City (Utah) (powdered carbon)	91	BOD = 13

Table 5 gives a partial list of direct physicochemical treatment plants either planned or currently under design.

Table 5

Physicochemical Treatment Plants

Site	Capacity	Status
Rocky River, Ohio	10	Bids Taken
Painesville, Ohio	5	Design
Cleveland, Ohio	10	Design
Fitchburg, Massachusetts	15	Design
Waterford, New York	2	Planned
Cortland, New York	10	Planned
Clay, New York	10	Planned
Niagara Falls, New York	60	Planned
Garland, Texas	30	Design
Owosso, Michigan	6	Design

8. Capital and Operating Costs

A detailed analysis of capital and operating costs has been carries out for a direct physicochemical treatment plant consisting of preliminary screening, coagulation, and adsorption for a treatment plant with a nominal design capacity of 10 MGD. The capital costs analysis is summarized in Table 6, and the annual operating costs analysis including amortization of 6% for 24 years, is summarized in Table 7.

TABLE 6

Estimated Capital Costs for Treatment of Raw Sewage by Clarification-Adsorption

Basis: 10 mgd Plant

EQUIPMENT, PIPING AND INSTRUMENTATION	TOTAL COST	
	Expanded Bed	Packed Bed
Pretreatment & Flocculation	$ 154,000	$ 154,000
Clarification & Sludge Handling	1,113,600	1,113,600
Adsorption System	551,000	617,600
Regeneration System	213,000	222,500
AUXILIARY FACILITIES		
Electric Power	51,000	80,000
Fuel Handling	20,000	20,000
Buildings & Structures	475,000	493,000
Roads, Walks, Fence	170,000	170,000
ACTIVATED CARBON	288,000	288,000
CONTINGENCY	377,000	390,000
TOTAL FIXED CAPITAL	$3,412,500	$3,548,700

TABLE 7

Estimated Annual Operating Costs for Treatment of Municipal Wastewater by Clarification and Adsorption in Expanded-Bed and Packed-Bed Adsorbers

Basis: 10 mgd

		Expanded	Packed
1.	Operating Labor*	$ 86,600	$ 86,600
2.	Maintenance Labor - 3% Plant Phys. Costs	53,600	55,600
3.	Maintenance Materials - 2% Plant Phys. Costs.	35,800	37,100
4.	Maintenance Supplies - 15% of 2 + 3	13,400	13,900
5.	Supervision - 15% of 1	13,000	13,000
6.	Payroll Overhead - 15% of 1 + 2	21,000	21,000
7.	General Overhead - 30% of 1 + 2 + 6	48,400	49,100
8.	Insurance - 1% of Plant Phys. Costs	17,900	18,600
9.	Carbon Makeup - 5% @ $.28/lb	27,500	27,500
10.	Lime Makeup - 25% @ $20/T	23,000	23,000
11.	Fuel - @ $0.50/MM Btu	75,000	75,000
12.	Power - $0.10/kwh	33,000	45,000
13.	Amortization - 24 years @ 6%	271,600	281,800
	Total Annual Cost	$719,800	$747,500
	Treatment Cost - ¢/1000 gal.	19.72	20.48

* 2 Shift Men + 2 day men @ $4.00 per hour

9. Sample Calculations and Typical Plots

A suggested flow sheet for the physicochemical treatment concept utilizing active carbon is given in Figure 4 and a possible plant lay-out in Figure 5. In this scheme, coagulant is added to the raw sewage, and flocculation takes place in a chamber which provides moderate agitation for an average detention time of 15 minutes. Clarification takes place in a sedimentation basin with an average detention time of two hours.

The clarified effluent is then passed through activated carbon adsorption units for removal of dissolved organics. The preferred mode of operation is an expanded bed, which permits the use of simple open-top concrete contacting basins and relatively trouble-free operation. The use of open tanks with trough-type overflows at the surface of the contacting basin provides a means of additional aeration of the wastewater during treatment, thus controlling anaerobic conditions such as those observed in the closed systems in the pilot plant. Two-stage contacting of the activated carbon is proposed to provide efficient utilization of adsorptive capacity by countercurrent movement of the activated carbon within the system. The plant lay-out for 10 mgd is based on installing five adsorption units of two stages each. When the granular carbon in the first stage of one unit is spent, that unit is to be taken off stream while the spent carbon is removed and regenerated in the multiple-hearth furnace system provided. During the time this unit is off-stream for the carbon regeneration, the other four units will run at 25% higher feed rate each. Upon completion of the carbon regeneration, the regenerated carbon will be

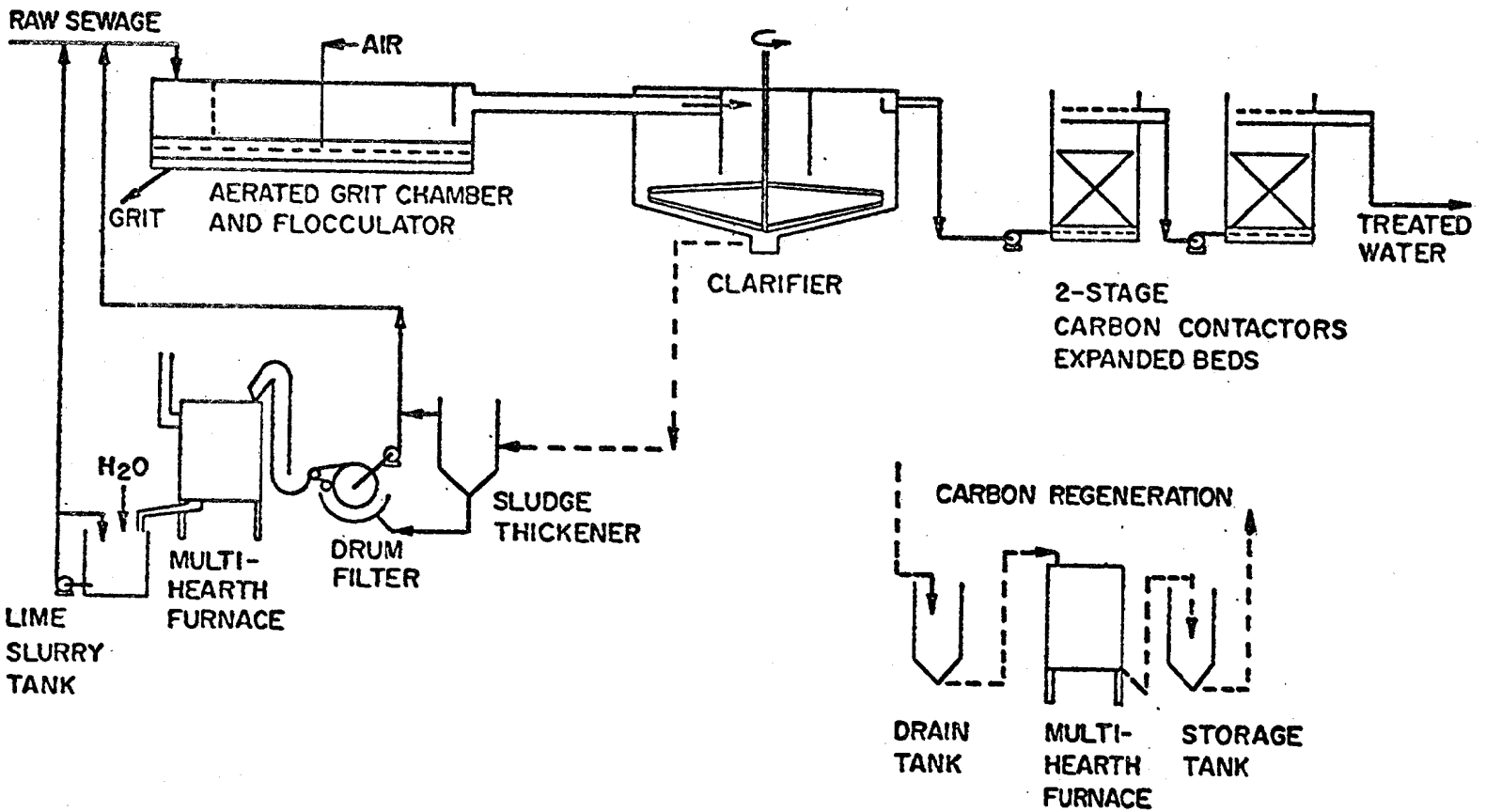


FIGURE 4. PROPOSED SCHEME OF TREATMENT OF RAW SEWAGE BY CHEMICAL CLARIFICATION AND ADSORPTION ON ACTIVATED CARBON

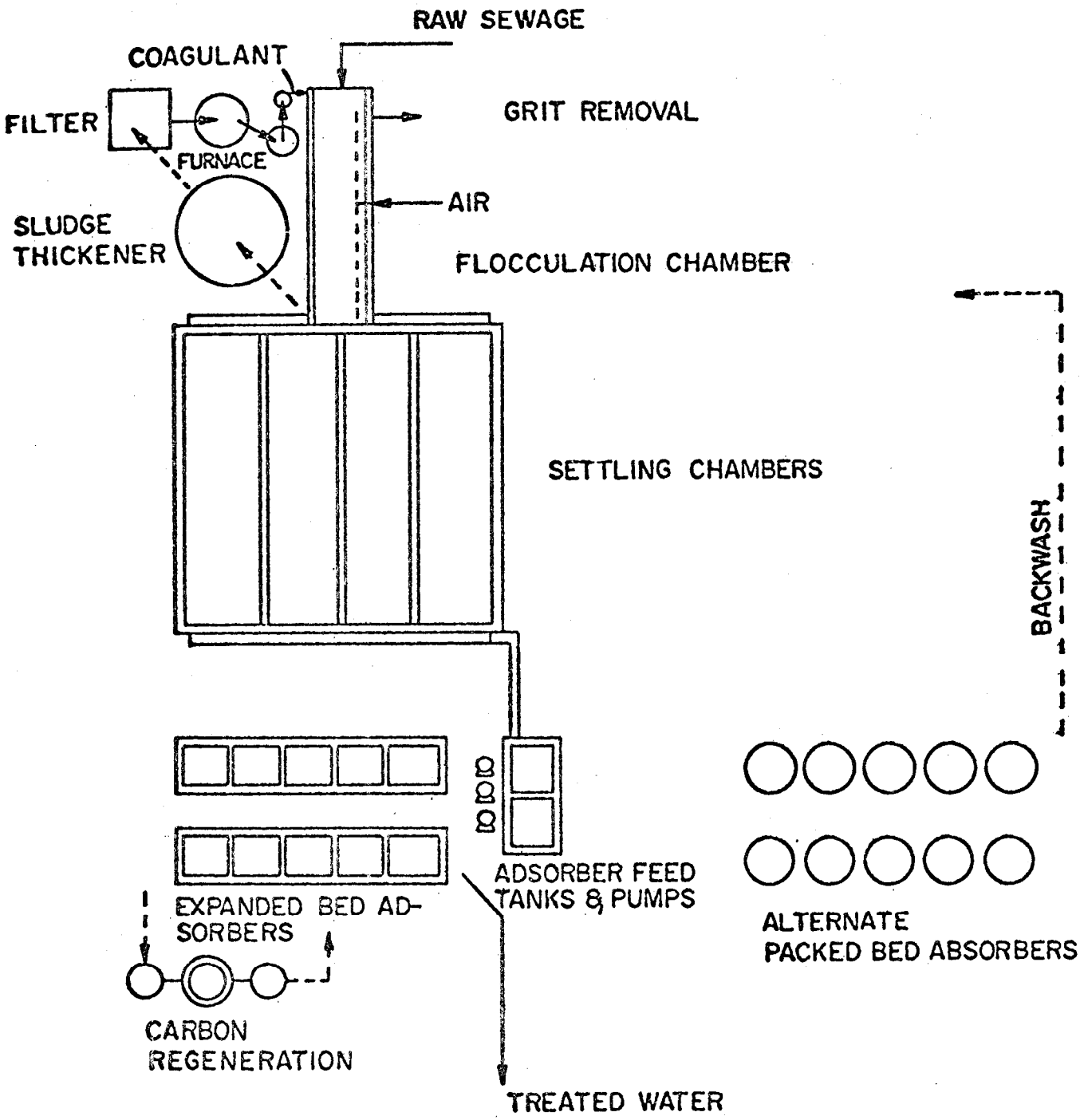


FIGURE 5. PROPOSED ARRANGEMENT OF PROCESSING UNITS

returned to the adsorber which will become the second stage of that unit; the former second stage with partially spent carbon becoming the first stage. Feed would then be evenly divided to the five units until another first stage carbon bed is spent.

The water resulting from the clarification and activated carbon treatment will enhance the quality of surface waters and with chlorination would be suitable for many uses. A final filtration treatment may be desirable to insure a crystal clear effluent for some reuse applications. This post-filtration would remove any suspended matter generated in the carbon columns which might otherwise be released into the final treated water.

10. Scale-up Factors

Scale-up of activated carbon adsorption systems is straightforward. No difficulties are anticipated for the design and implementation of carbon adsorption systems for any size waste treatment facility. Restrictions on vessel size are involved with pressure packed-bed adsorbers, but these are not recommended for general use.

11. Process Dependability

The stability of operation provided by a treatment system based on physical and chemical technology is perhaps the most important single advantage of physicochemical treatment systems employing adsorption. Biological systems are notoriously sensitive to changes in environmental conditions. If a toxic material gains even temporary entrance to a biological plant, or a hydraulic peak occurs not only will the efficacy of the plant drop off but recovery may take from several days to several weeks. In a physicochemical plant the filtration system backs up the clarifier and the carbon system backs up the filter; thus upsets are unlikely. Further, it can be expected that an immediate recovery of the plant will take place once the source of upset is eliminated. This inherent stability of performance is also reflected in greater design and operational flexibility. Whole sections of a physicochemical plant can be cut in or out of the process stream as required, and a temporary overload can be absorbed with little effect. The major advantages of a physicochemical system are given in Table 8.

Table 8

Advantages of Physicochemical vs. Conventional Secondary Treatment

1. Less Area Requirement - 1/2 to 1/4
2. Lower Sensitivity to Diurnal Variation
3. Not Affected by Toxic Substances
4. Potential for Significant Heavy Metal Removal
5. Superior Removal of P Compounds
6. Greater Flexibility in Design and Operation
7. Superior Organic Removal

12. Matrix Summary

Parameter of Concern	Target Year 1975	1990	2020
Scale-up	Direct	Direct	Direct
Dependability over time	High	High	High
Flexibility (relative to other processes)	High	High	High
Tolerance to Influent Variability	High	High	High
Ancillary Processes Required	Few	Few	Few
Costs	Favorable	Favorable	Favorable
Process Efficiency	95%	95%	90%
Operational Difficulty	No Major Difficulty	NMD	NMD

REFERENCES

Beebe, R.L. and Stevens, J.I., "Activated Carbon System for Wastewater Renovation", Water and Waste Engineering, 4, (1), p. 43 (1967).

Davies, D.S. and Kaplan, R.A., "Activated Carbon Eliminates Organics", Chemical Engineering Progress, 60, (12), p. 46 (1964).

Hager, D., and Reilly, P., "Clarification-Adsorption in the Treatment of Municipal and Industiral Wastewater", Journal Water Pollution Control Federation, 42, 5, 1970.

Johnson, R.L., Lowes, F.J., Jr., Smith, R.M. and Powers, T.J., Evaluation of the Use of Activated Carbons and Chemical Regenerants in Treatment of Wastewater, 999-WP-13, U. S. Dept. Health, Education, and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1964.

Joyce, R. S. and Sukenik, V.A.

1964: Feasibility of Granular Activated-Carbon Adsorption for Wastewater Renovation, 999-WP-12, W. S. Dept. Health, Education, and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1964.

1965: Feasibility of Granular Activated-Carbon Adsorption for Wastewater Renovation 2, 999-WP-28, U. S. Dept. Health, Education and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1965.

Keinath, T.M. and Weber, W.J., Jr.,"A Predictive Model for the Design of Fluid-Bed Adsorbers", Journal of the Water Pollution Control Federation, 40, (5), p. 741 (1968).

Morris, J.C. and Weber, W.J., Jr., Preliminary Appraisal of Advanced Waste Treatment Processes, TRW 62-24, U.S. Dept. Health, Education, and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1962.

Parkhurst, J.D., Dryden, F.D., McDermott, G.N., and English, J., "Pomona Activated Carbon Pilot Plant", Journal of the Water Pollution Control Federation, 39, (10), p. R70 (1967).

PHS

1962: Summary Report, Advanced Waste Treatment Research Program, June 1960-December 1961, TRW 62-9, U.S. Dept. Health, Education and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1962.

1965: Summary-Report, Advanced Waste Treatment Research Program, January 1962-June 1964, 999-WP-24, U.S. Dept. Health, Education and Welfare, R.A. Taft Sanitary Engineering Center, Cincinnati, Ohio, 1965.

Schlecta, A.F. and Culp, G.L., "Water Reclamation Studies at the South Tahoe Public Utility District", Journal of the Water Pollution Control Federation, 39, (5), p 787 (1967).

Stevens, D.B. and Peters, J., "Long Island Recharge Studies", Journal of the Water Pollution Control Federation, 38, (12), p 2009 (1966).

Weber, W.J., Jr. and Morris, J.C., "Kinetics of Adsorption on Carbon from Solution", Journal of the Sanitary Engineering Division, ASCE, 89, SA2, p 31 (1963); Closure: Ibid., 89, SA6, p 53 (1963).

Weber, W.J., Jr. and Atkins, P.F., Jr., "Physical Separation Methods for Advanced Treatment of Wastewater" Proceedings of the National Symposium on Quality Standards for Natural Waters, ASCE, University of Michigan, Ann Arbor, Michigan, p 137, 1966.

Weber, W.J., Jr., "Fluid-Carbon Columns for Sorption of Persistent Organic Pollutants", Advances in Water Pollution Control Research, Proceedings of the Third International Conference of Water Pollution Research, Water Pollution Control Federation, Washington, D.C. Vol. I, p 253, 1967.

Weber, W. J., Jr., Hopkins, C. B. and Bloom, R., "A Comparison of Expanded-Bed and Packed-Bed Adsorption Systems", R. A. Taft Water Research Center Report No. TWRC-2, FWQA, Cincinnati, Ohio, 1968.

Weber, W. J., Jr., Hopkins, C. B. and Bloom, R., Jr., "Granular Carbon Treatment of Raw Sewage", Water Pollution Control Research Series ORD-17050 DAL 05/70, FWQA, Cincinnati, Ohio, 1970.

Weber, W. J., Jr., Hopkins, C. B. and Bloom, R., Jr., "Physicochemical Treatment of Wastewater", Journal Water Pollution Control Federation, 42, 1, 1970.

Report to
North Atlantic Division
Corps of Engineers
U. S. Army

on

WASTEWATER TREATMENT BY COAGULATION IN THE MERRIMACK RIVER BASIN

by

Mark W. Tenney
TEN-ECH, Inc.
52580 Dixie Way, North
South Bend, Indiana 46637

June 1971

TABLE OF CONTENTS

		Page
1.	Definition of the Problem	1-5
2.	Concept and Theory - Summary	6-19
3.	Applicability to the Systems in Question	20-25
4.	Design Criteria	26-31
5.	Pre-Treatments Required	32-33
6.	Effluent Qualities	34-36
7.	Experience on Process to Date	37-38
8.	Capital and Operating Costs	39-41
9.	Sample Calculations and Typical Plots	42-44
10.	Scale Up Factors	45
11.	Process Dependability	46
12.	Matrix Summary	47

I. DEFINITION OF THE PROBLEM

This report on coagulation technology is one of a series of reports prepared by a group of water consultants retained by the Corps of Engineers (North Atlantic Region) to prepare information on advanced wastewater treatment methods applicable to their Merrimack River Basin study. The Merrimack River, which is one of the major rivers in the United States, eminates in upper New Hampshire, flows essentially in a southerly direction into central Massachusetts where it turns abruptly north-east and ultimately discharges into the Atlantic Ocean north of Boston, Massachusetts. The river is approximately 116 miles in length, has seven principal tributaries and encompasses a drainage area of approximately 5,000 square miles.

The Merrimack River Basin study is one of a number of wastewater management projects currently under study by the Corps of Engineers. The purpose of these studies is to focus the planning capability and experience of the Corps of Engineers on the problems of total water management and to seek out new technologies consistent with contemporary environmental objectives. In this regard, coagulation is one of a number of unit processes that can have application in achieving maximum improvement of water quality in this basin.

Although the overall study encompasses six areas for immediate action, namely: the Winnipesaukee River, the area between Concord and Manchester, the area at Nashua, the Lowell-Lawrence-Haverhill area, the Fitchburg area,

and the metropolitan area of Boston (limited to major waste discharge points in the harbor and immediate bay areas); three separate regions have been given priority for this studv:

1. The Metropolitan Boston Area,
2. The Lowell-Lawrence-Haverhill Area, and
3. The Winnipesaukee River Area.

In preliminary discussions, the major objectives of the waste treatment in these areas were established to be as follows:

1. reduction of BOD_5 to 2 mg/l,
2. reduction of Suspended Solids to 2 mg/l,
3. reduction of COD to 5 mg/l,
4. reduction of total Nitrogen to less than 1 mg/l as N,
5. reduction of phosphorous to less than 1 mg/l as P, and
6. reduction of total dissolved solids to 100 mg/l.

The above criteria were selected as the primary objectives of any treatment scheme to be specified. Other water quality parameters noted as being important included pH, color, turbidity, grease, alkalinity, hardness, metals, MPN, viruses, chlorides, toxins, and sulfate.

Two types of wastewater treatment schemes were proposed for consideration to achieve the above water quality requirements.

Scheme I. This scheme consists of conventional biological treatment, followed by biological nitrification with concurrent coagulation, sedimentation, adsorption (with denitrification), filtration, ion exchange, and membrane concentration of the brine solution from ion exchange prior to ocean disposal. Sludge treatment will consist of thickening, vacuum filtration and incineration. Figure I shows a block diagram for this treatment scheme.

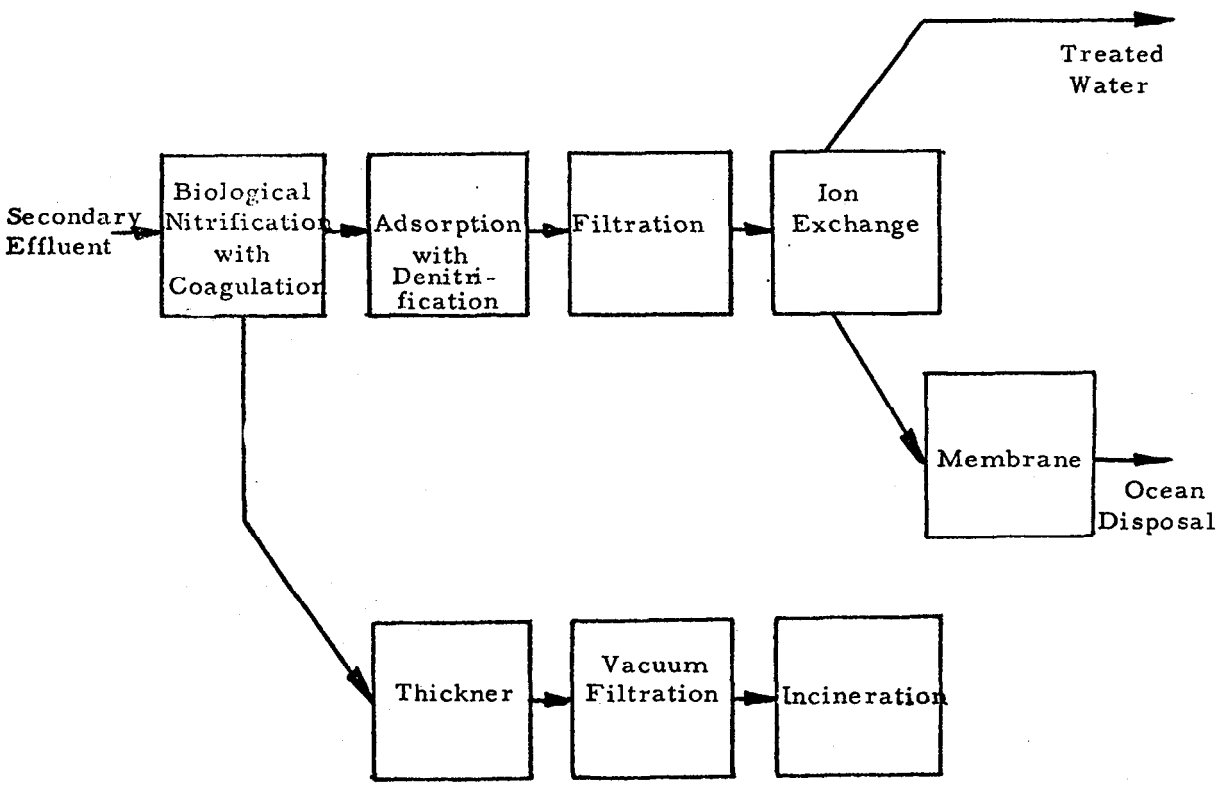


Figure 1: Block Diagram of Proposed Treatment Scheme I. (Biological-Tertiary Treatment)

Scheme II. This scheme is a physical-chemical treatment process consisting of coagulation, adsorption (with break point chlorination for ammonia nitrogen removal), filtration, and ion exchange with membrane concentration of the brine solution from ion exchange for ocean disposal. The sludge treatment scheme involves thickening, vacuum filtration, and incineration. Figure 2 shows a block diagram for this treatment scheme.

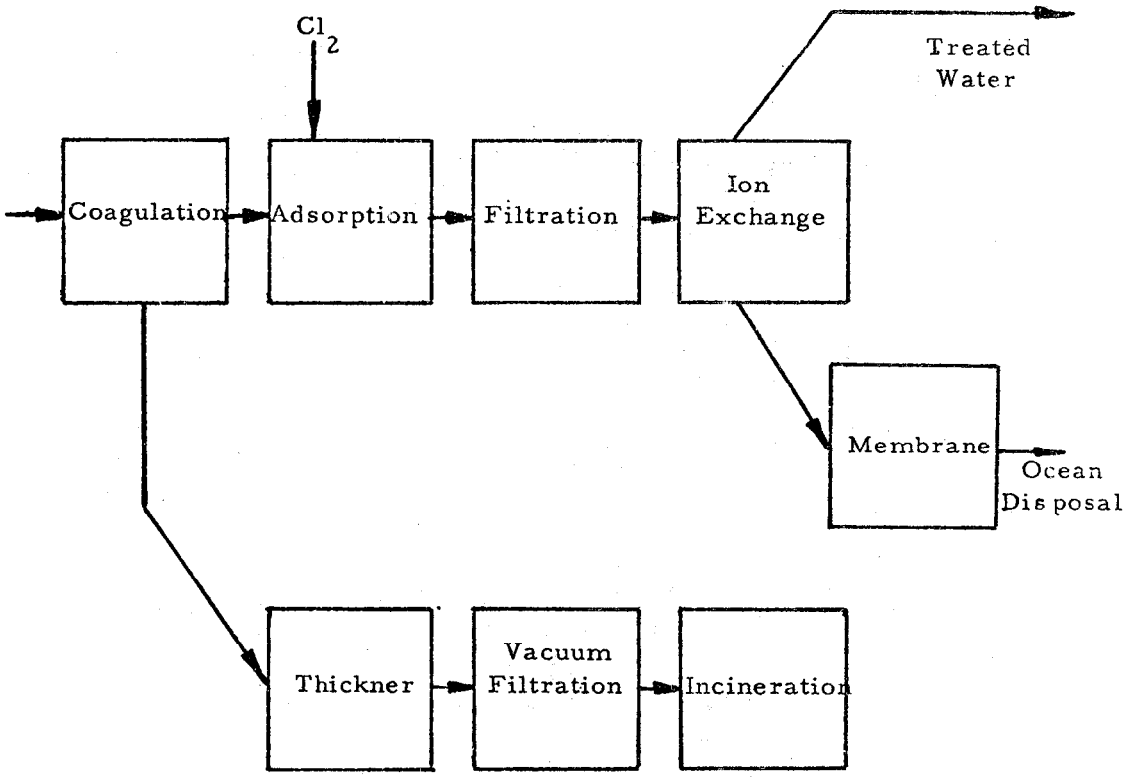


Figure 2: Block Diagram of Proposed Treatment Scheme II. (Physical-Chemical Treatment)

Both treatment schemes will be considered for each of the three priority regions designated in the river basin. This report describes the use of coagulation in the two proposed treatment schemes. The report follows an outline (see Table of Contents) agreed upon in a preliminary meeting of consultants and representatives of the Corps of Engineers in Chicago, Illinois on April 29, 1971.

2. CONCEPT AND THEORY - SUMMARY

The coagulation process can be utilized in the treatment of wastewaters to remove small sized suspended impurities that otherwise would not be removed by plain sedimentation. Typical impurities that are effectively removed by coagulation include suspended clays, silts, bacteria, algae, viruses, etc. Nominally these impurities can be classified as colloidal. They tend to form stable suspensions principally because they inherently possess a charged surface which, among other factors, precludes their agglomeration to larger particles capable of obtaining effective settling velocities. Under certain conditions this process will also remove certain dissolved impurities; phosphatic pollutants and high molecular weight organic materials (e.g., color-causing compounds, refractory organic materials, etc.) are frequently removed in this manner.

The basic mechanism involved in coagulation is to agglomerate the suspended and/or dissolved impurities into larger particles or packets which subsequently can be effectively separated from the water by such solids-liquid separation processes as sedimentation, granular filtration, centrifugation, etc. The process characteristically consists of three phases:

1. chemical dosing
2. agitation
3. separation

A chemical coagulant is added in a chemical dosing tank that is mixed rapidly in order to achieve rapid and uniform coagulant dispersion in short hydraulic detention times (characteristically in the neighborhood of sixty seconds). After chemical dosing the water enters an agitation tank where agitation is generally provided by slowly rotating paddle type devices designed to agglomerate the individual particles; the hydraulic retention time in the agitation tank is generally in the range of 15-30 minutes after which the agglomerated suspension can be separated from the water phase by one or more of the processes mentioned above.

Chemical coagulants effective in wastewater treatment generally fall in one of two groups: 1.) inorganic polyvalent metal ions (iron, aluminum salts) or synthetic organic polyelectrolytes (anionic, cationic, or non-ionic). Selection of a coagulant is governed by the characteristics of the wastewater and the extent of treatment desired. Determination of the requisite dosage of the selected coagulant is singularly critical as insufficient concentrations will not achieve effective agglomeration, whereas, in many cases, excess concentrations will preclude agglomeration. The requisite coagulant dose is influenced by a number of parameters, some of the most significant of which can include:

1. pH
2. alkalinity
3. impurity concentration
4. anion-cation concentration and composition

Coagulation Theory

A typical coagulation (or flocculation) curve is shown in Figure 3. The data shown in this figure indicate the effect of increasing concentrations of a cationic synthetic organic polyelectrolyte on the flocculation of dispersed microorganisms (principally bacteria). Note that with low polyelectrolyte concentrations no flocculation occurs until a critical concentration of polyelectrolyte is reached - - - at which point flocculation begins. Increasing the polyelectrolyte concentration above this critical value improves the flocculation until an optimum concentration is reached. Finally, it can be observed that concentrations of polyelectrolyte greater than the optimum ultimately cause redispersion of the system.

Figure 4 indicates a similar flocculation curve of microorganisms (principally bacteria) with an inorganic flocculant (alum). No re-dispersion is observed in this case due to solubility restrictions of aluminum. Also shown on this figure is the concurrent removal of phosphate. The theory of the precipitation of phosphate with polyvalent metal ions is discussed later in this section; the preferential precipitation of soluble phosphatic species by the metal ion prior to coagulation of the suspended microorganisms should be noted and is typical for such systems.

The shape of the curves in Figures 3 and 4 are typical for either a flocculation or coagulation phenomenon. As alluded to above, the difference between the two phenomena depends on the mechanism of the agglomeration and

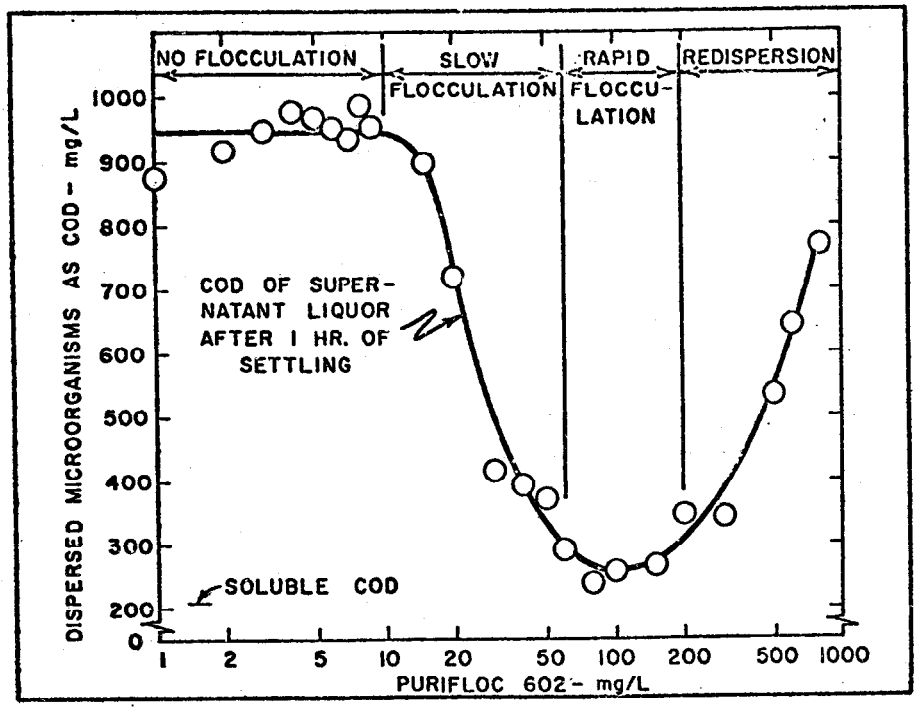


Figure 3: Typical Flocculation Curve for Dispersed Microorganisms with Cationic Polyelectrolyte

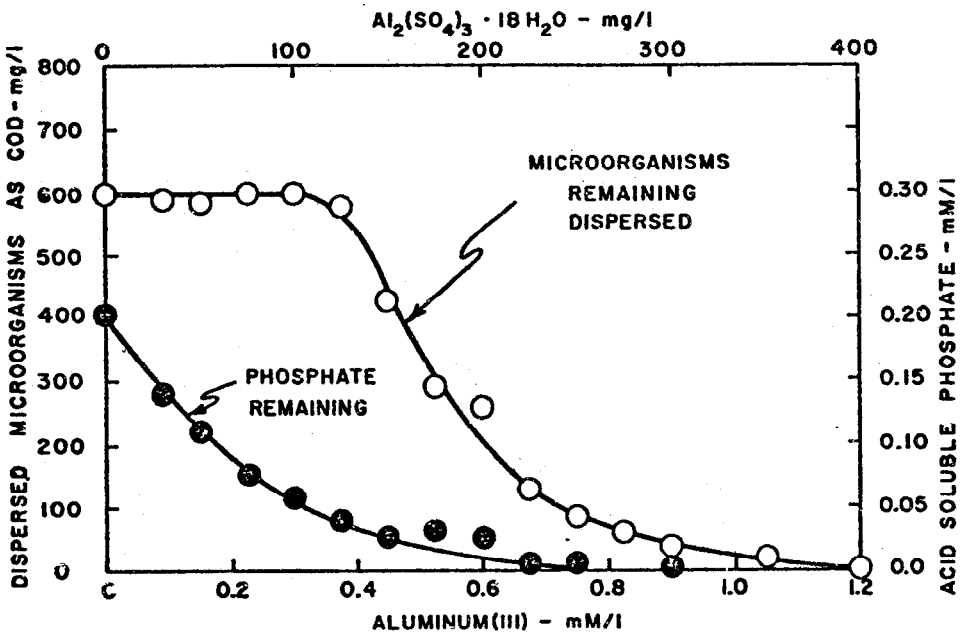


Figure 4: Typical Flocculation Curve for Dispersed Microorganisms with Aluminum in the Presence of Ortho-Phosphate

ultimately dictates the characteristics of the aggregate formed (i. e., loose or tight). This distinction is illustrated by Figure 5A which shows the flocculation of algae with a cationic synthetic organic polyelectrolyte. In coagulation (Figure 5B), the coagulant serves to reduce the surface charge on the colloid below a given millivolt level (plus or minus) such that it can be overcome by the inherent kinetic energy of the particle and ultimately be held together by the London van der Walls forces of attraction. Thus coagulation occurs between a limiting plus and minus millivolt level (e. g., $\pm$ 15 mv) and at greater absolute values than this value no aggregation occurs. In flocculation (Figures 5C and 5D) the flocculant serves to attach itself to the surface of the colloid by either chemical or electrostatic forces and subsequently bridge and attach its extended segments to exposed surface sites on other colloid particles. Under these conditions, initially insufficient polymer will be present to withstand the shearing forces of agitation, whereas in the presence of excessive flocculant concentrations an insufficient number of surface sites will be available for attachment of bridging polymeric species and thus in both instances no aggregation will result. Consequently under these conditions the extent of polymer coverage of the colloid surface becomes the controlling mechanism for flocculation. Incidently, the data in Figure 5 represent a flocculation phenomenon and the close agreement of the point of zero surface charge (isoelectric point - see Figure 5B) with optimum flocculation is coincidental.

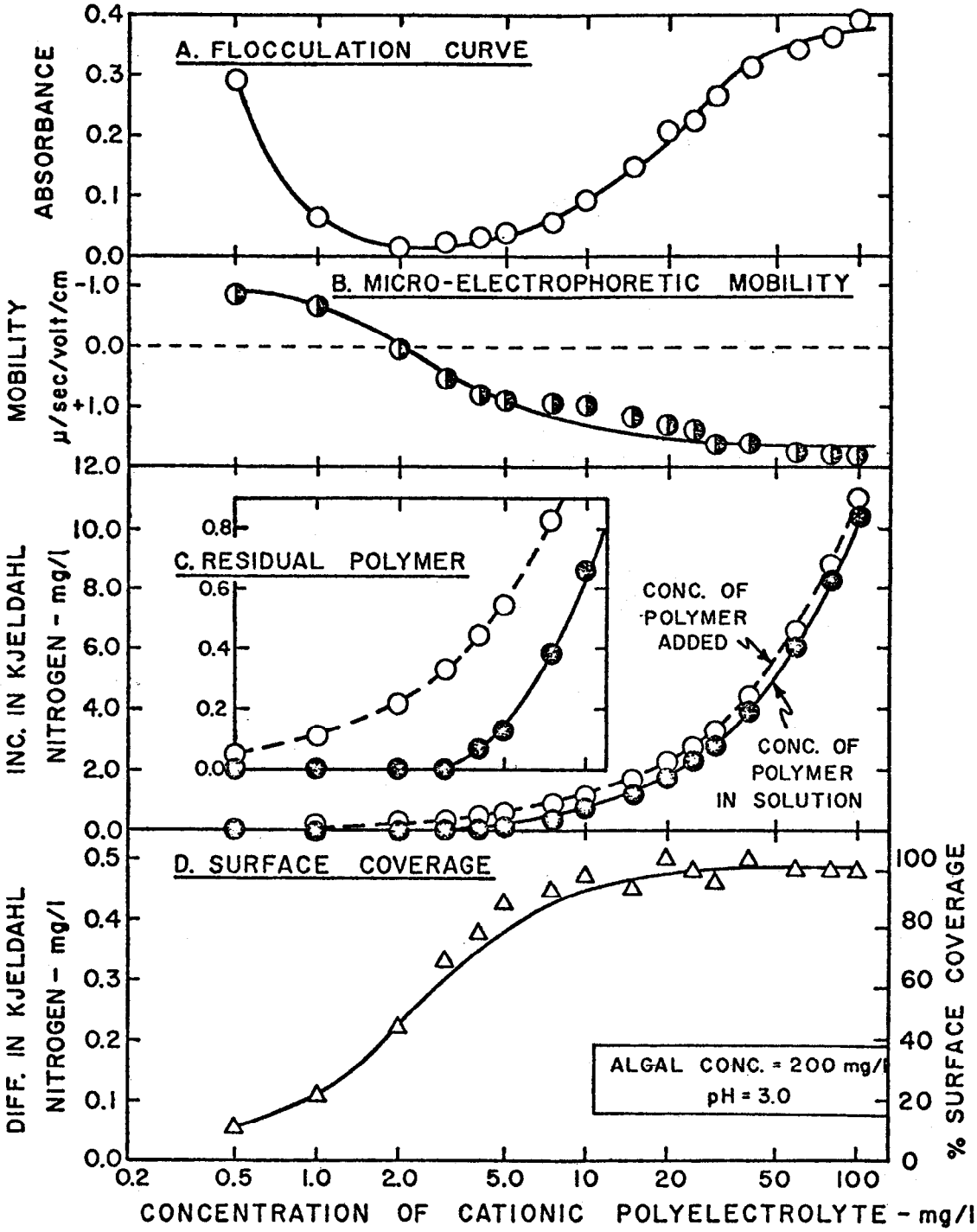


Figure 5: Flocculation of Dispersed Algae with Synthetic Organic Polyelectrolyte

Major Factors which Influence Coagulation (Flocculation)

pH. Hydrogen ion can affect the flocculation in various ways: it influences the charge characteristics of both the colloid and the coagulant. In the case of synthetic organic polyelectrolyte, it determines the extent of coiling and influences the diffusion characteristics of the polymer. The pH dependence of flocculation by metal ions is instructive (Figures 6 and 7). The optimum pH range for effective flocculation is around pH 5 to 6 for aluminum (III) (Figure 6), and around pH 4.5 to 5.5 for iron (III) (Figure 7). Within this pH range, all the aluminum (III) and iron (III) ions undergo rapid hydrolysis and form linear hydroxopolymers. At pH values lower that this optimum range, where the free metal ions Al^{+3} and Fe^{+3} are prevalent, and in the neutral and alkaline pH range, where hydrolysis leads immediately to hydrous metal oxides, ready flocculation of microorganisms does not occur. In the acid pH range ($pH < 4$) where the charge of the microbial surface is less negative than in nearly neutral solutions and thus supposedly more amenable to coagulation, free multivalent ions of iron (III) and aluminum (III) at substantial concentrations, far above the critical concentration necessary for the coagulation of lyophobic colloids, are not capable of coagulating microbial cells.

Hydrolysis products alone, and not the free metal cations or hydrous metal oxides, produce agglomeration of colloids. The aggregation produced by aluminum (III) or iron (III) salts is caused by the interaction of linear polymers, resulting from the hydrolysis of these salts, with the dispersed

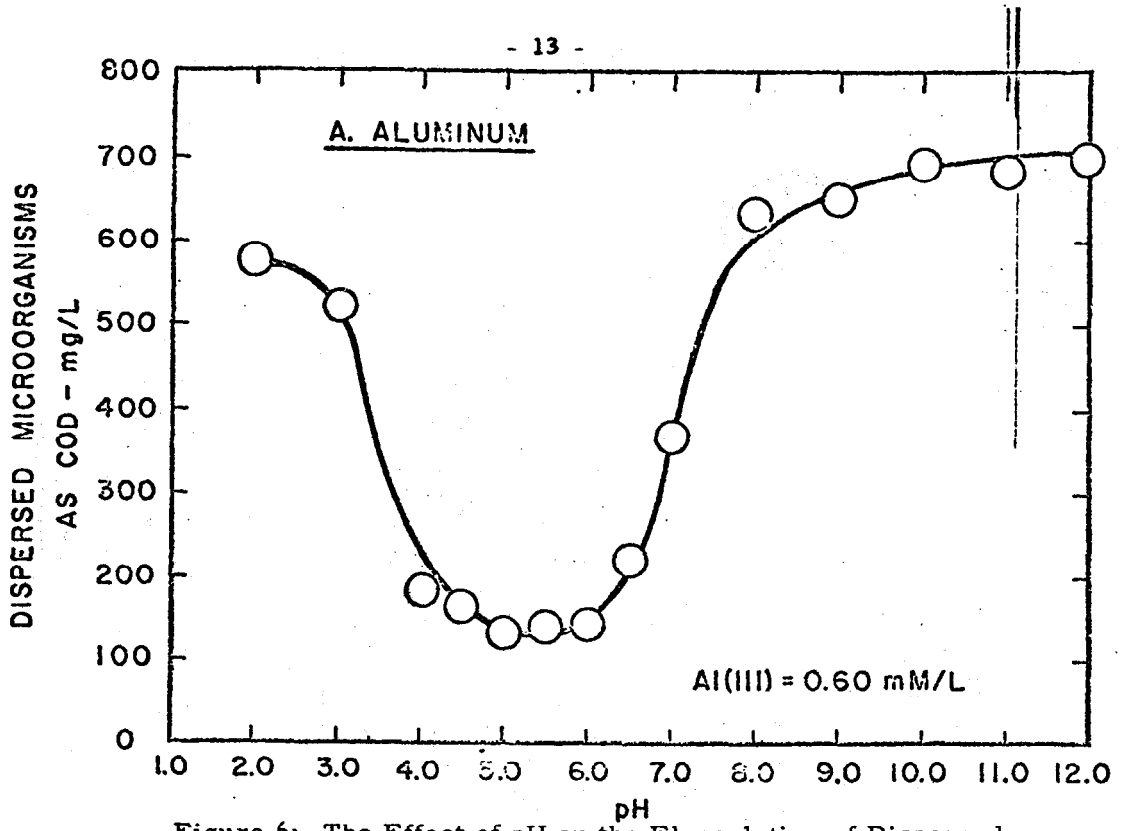


Figure 6: The Effect of pH on the Flocculation of Dispersed Bacteria with Aluminum.

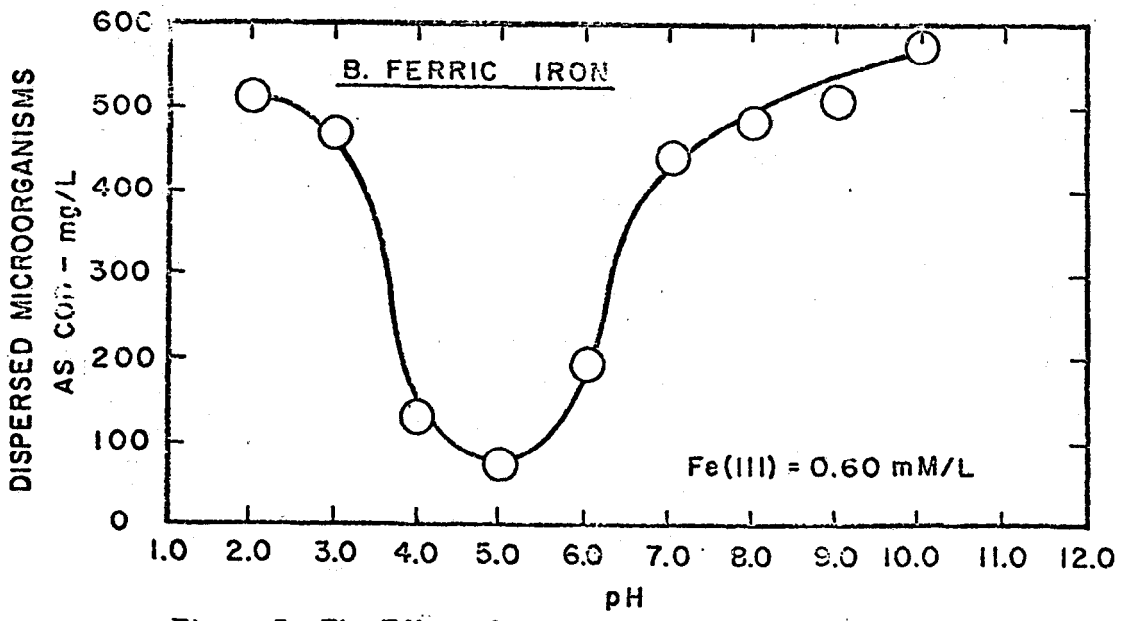


Figure 7: The Effect of pH on the Flocculation of Dispersed Bacteria with Iron

colloidal particles. The metal ions at the end of these short chain polymers will be bonded predominantly by chemical forces onto the ionogenic groups of the colloid surface.

Stoichiometry. Values of the amount of polymer necessary to flocculate 50 percent of the colloids (microorganisms) under controlled conditions of pH, agitation and temperature are presented in Figure 8. A linear relationship between the amount of flocculant and the number of microorganisms, i. e., total specific surface area exposed is obtained in each case. Qualitatively, this "stiochiometric" relationship can be explained by assuming that a certain and approximately constant fraction of the surface of the cells has to become covered with extended polymer segments.

Presumably, almost all polymer molecules that have been added to the solution become adsorbed at the microbial surfaces. It appears from these results that the amount of polyelectrolyte required to cause flocculation is of the order of 20μg/mg of cell (dry weight). This amount appears to be approximately one-half of that which can ultimately be adsorbed at the cell surface (Figure 5D). The addition of an excess of polyelectrolyte to the colloidal suspension thus leads to a partial redispersion of the organisms (Figure 3). As soon as too large a fraction of the cell surface becomes covered with polyelectrolyte, the surfaces which are now positively charged start to repel each other.

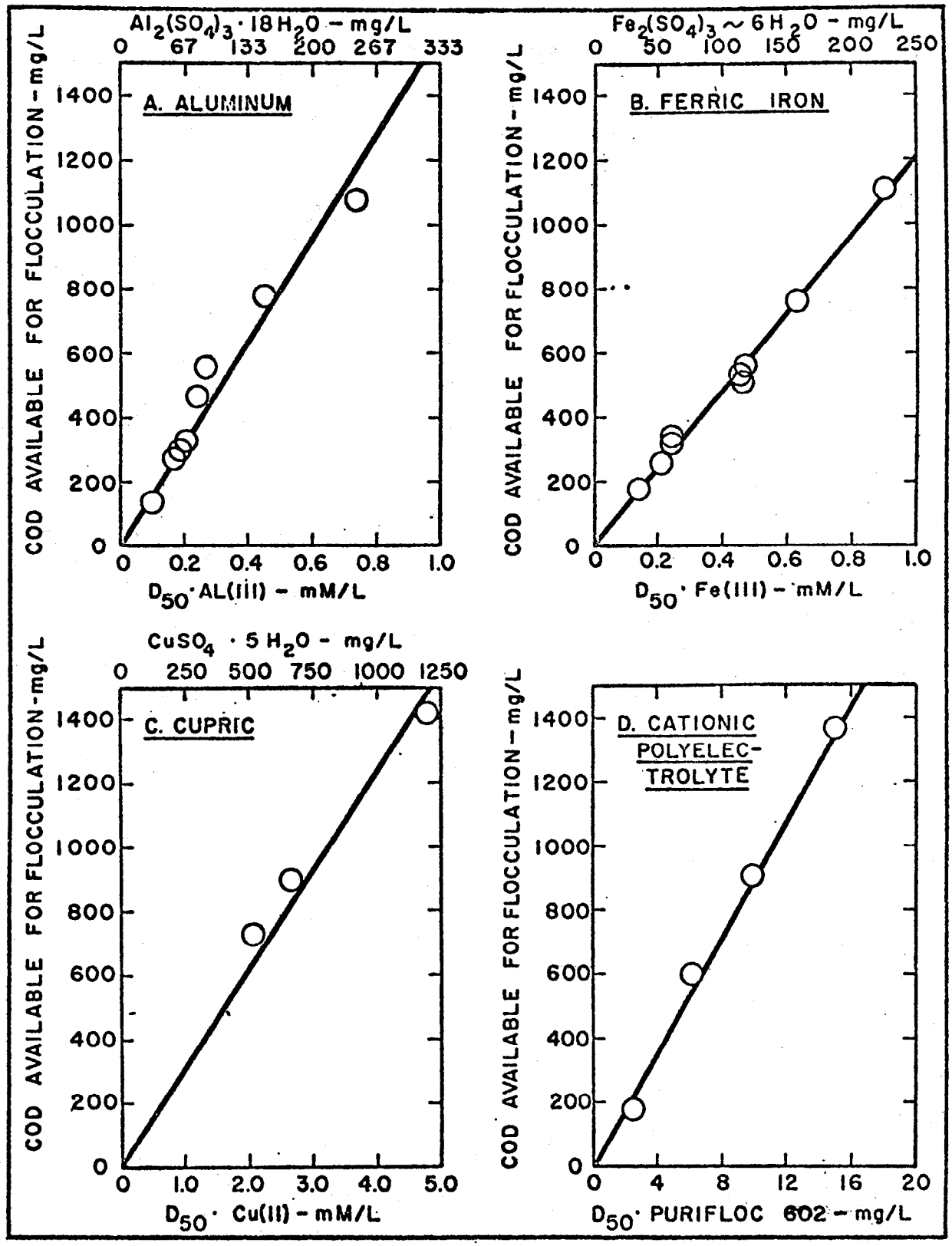


Figure 8: Influence of Increasing Colloid Concentration (Microorganism) on the D_{50} Flocculant Dose

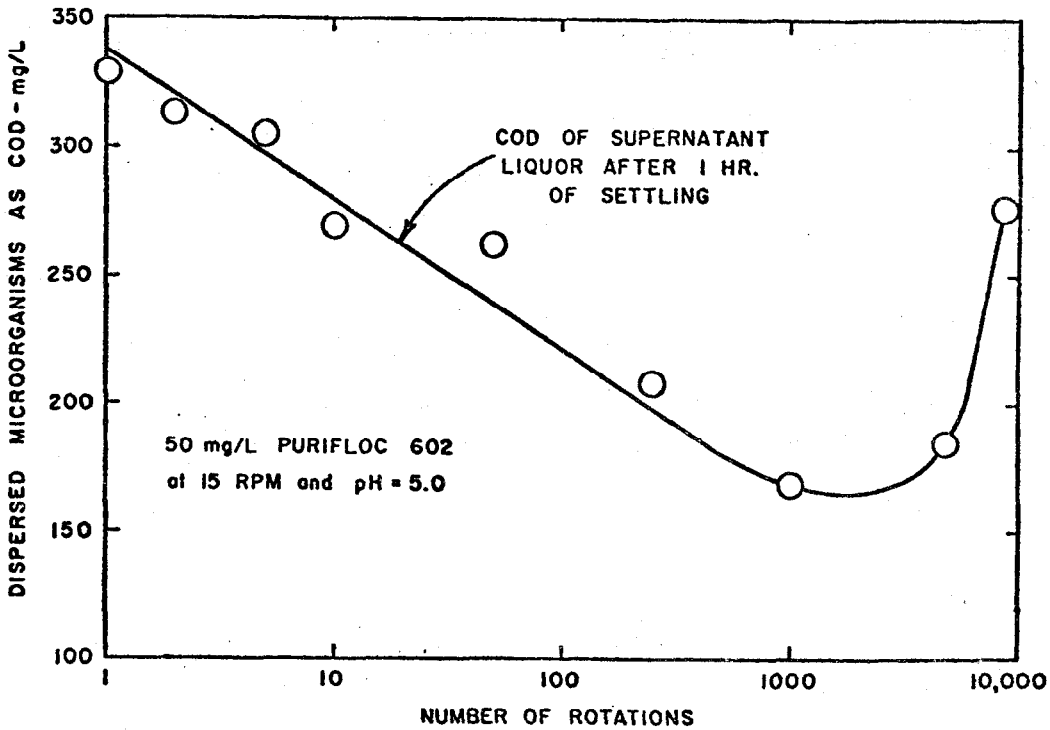


Figure 9: The Effect of Agitation on the Flocculation of Bio-colloids

Agitation. Figure 9 indicates the effect of agitation on flocculation. With a given colloid-flocculant suspension, the extent of the flocculant adsorption increases with prolonged agitation. Hence, optimum flocculation is obtained only for a certain time span and if the suspension is agitated for a longer time period it redisperses, preseumably because of more extensive adsorption of polymer segments on the cell surface.

Chemical Precipitation of Phosphate. Many metal ions are particularly effective in precipitating phosphate from solution. Due to their relatively low cost and general availability, the three most frequently used metals for phosphate precipitation are calcium, iron, and aluminum (cost figures are presented in Section 3, Table 2). An examination of Table 1 reveals that all three of these metals form relatively insoluble precipitates with phosphate.

Table 1: Equilibrium Constants Related to Phosphate Precipitation with Metal Ions.

Reaction	Log equilibrium constant, 25°C
$Fe^{+3} + PO_4^{-3} = FePO_4(s)$	23
$3Fe^{+2} + PO_4^{-3} = Fe_3(PO_4)_2(s)$	30
$Al^{+3} + PO_4^{-3} = AlPO_4(s)$	21
$Ca^{+2} + 2H_2PO_4^{-} = Ca(H_2PO_4)_2(s)$	1
$Ca^{+2} + HPO_4^{-} = CaHPO_4(s)$	6
$10Ca^{+2} + 6PO_4^{-3} + 2OH = Ca_{10}(OH)_2(PO_4)_6(s)$	90

The tendency of aluminum and iron to hydrolyze in aqueous solution creates a competition between the hydroxide and phosphate ions for the coordination sphere of the metal. Thus the efficiency of phosphate removal is dependent upon the relative concentrations of these two anions in solution and is consequently pH dependent. A decrease in pH, or, more precisely, (OH^-), favors precipitation of the metal phosphate. However, as the solubility of the metal phosphates increases with decreasing pH, an optimum pH exists for the removal of phosphate with metal ion precipitants. When calcium is used as a precipitant, the competition for calcium is predominatly between the phosphate and carbonate anions and, again, phosphate removal is dependent upon the relative concentrations of the anions present, and upon pH. Hydroxylapatite, $Ca_{10}(OH)_2(PO_4)_6$, is the most stable calcium phosphate solid phase as shown in Table 1.

Under favorable pH conditions, the requisite concentration of Al(III) for the removal of phosphate closely approximates the one to one Al to PO_4 ratio which should theoretically exist in accordance with the reaction shown in Table 1. This is shown by the "filtered" line of Figure 10. The effect of pH on phosphate removal by Al(III) is shown in Figure 11. The filtered line of Figure 10 was obtained by filtering the insoluble $AlPO_4$ from solution. Operationally, where filtration may be impractical, excess Al(III) must be added to effect useful phosphate removal as shown by the settled line in Figure 10. Although the phosphate has been rendered insoluble by the aluminum, additional Al(III) is required as a coagulant. It must be realized that chemical precipitation of phosphate is a two stage process: (1) the phosphate must be made insoluble through the use of metal ions; and (2) the precipitate must be removed from solution by sedimentation. The addition of various polyelectrolytes, or the addition of excess metal ions to form metal hydroxides serve identical purposes in bringing about agglomeration of the insoluble metal phosphate particles into a settleable mass.

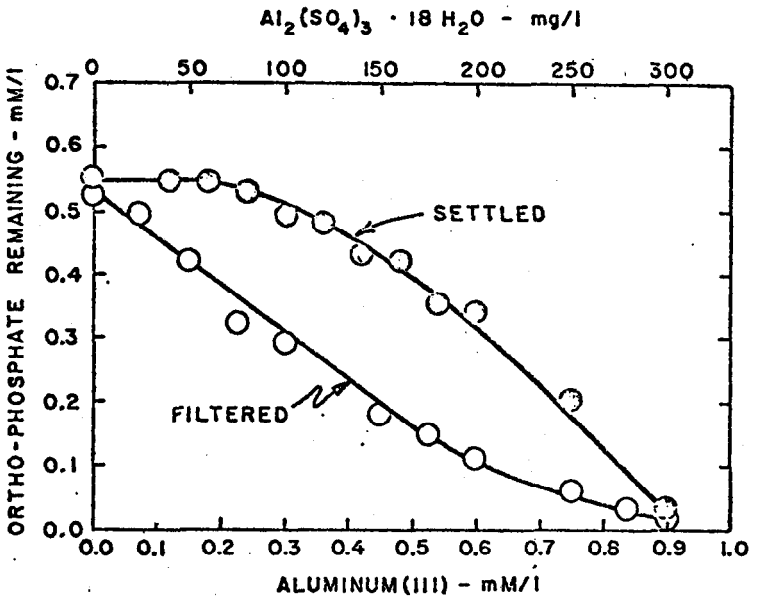


Figure 10: Removal of ortho-Phosphate with Aluminum.

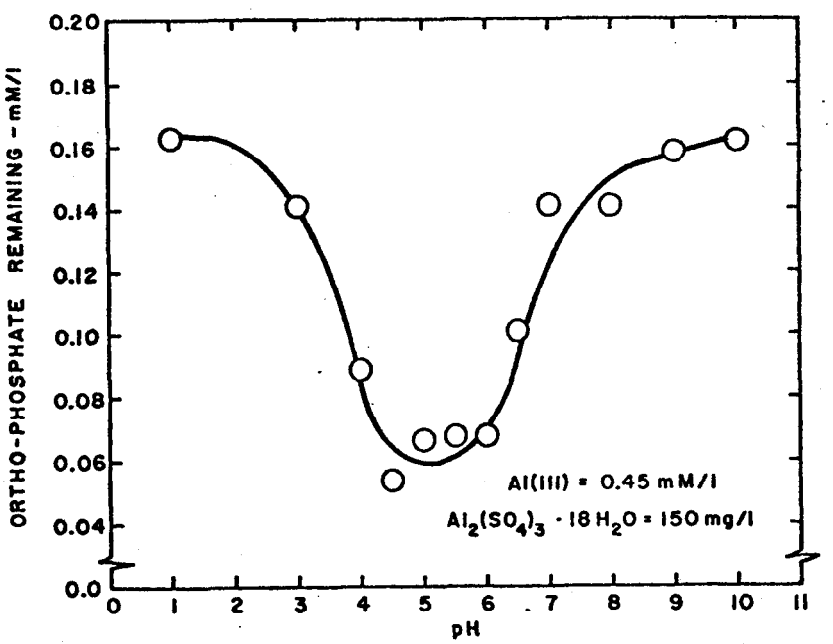


Figure 11: The Effect of pH on the Removal of ortho-Phosphate with Aluminum.

3. APPLICABILITY TO THE SYSTEMS IN QUESTION

Coagulation (flocculation) processes have been successfully employed in water treatment systems for many years. These processes have been selected routinely to effectively remove color and turbidity causing substances from water. An excellent review of the applications and associated theories of coagulation has just been completed (AWWA Coagulation Committee, 1971); another good review was published earlier by O'Melia (O'Melia, 1969). Over the years this process has proven itself to be extremely adaptable and flexible to changes in water quality and by proper adjustment of coagulant doses, pH, and agitation, it is possible to achieve a treated water of constantly high purity. In addition to its dependability, the process has proven itself to be relatively simple to control and amenable to monitoring by instrumentation.

The concept of employing coagulation processes in wastewater treatment plants is by no means new. In fact, the composition of the impurities in wastewater is such that they are for the most part much more amenable to removal by coagulation techniques then the corresponding impurities encountered in water treatment. Coagulation, for example, has been recognized for many years as being capable of appreciably improving the effectiveness of primary treatment of wastewater and during the last few years the complementary nature of coagulation and secondary treatment of wastewater has been studied and successfully reported (Tenney and Stumm, 1965) (Barth

et al., 1969). Due to such success and the fundemental soundness and simplicity of this process, its use as a separate tertiary treatment process has been widely advocated (Slechta and Culp, 1967)(Nesbitt, 1969) (Hall and Engelbrecht, 1969) (Little, 1970) (EPA Symposium, 1971) as has its use as a replacement for biological treatment and tertiary treatment via physical-chemical treatment (Weber, et al., 1970) (Zuckerman and Molof, 1970).

Coagulation can have direct application in both of the treatment schemes proposed for the various priority areas of consideration in the Merrimack basin. Flow sheets indicating the relative position of coagulation in the two proposed treatment schemes are shown on pages 3 and 4. By positioning and use of the coagulation process as shown, removal of appreciable fractions of the following pollutants can be expected (see section 6 for estimated influent and effluent values):

A. Scheme I: Biological-Tertiary Treatment

1. Suspended Solids. Coagulation will remove large fractions of the suspended and colloidal impurities (organic and inorganic) that were not removed by primary and secondary treatment - - - including unsettled (or dispersed) activated sludge microorganisms as well as algae and viruses.
2. Phosphatic Pollutants. If inorganic coagulants are selected (i. e., aluminum, iron, calcium), dissolved species of both ortho and poly phosphate can be effectively removed from wastewaters.

3. Organic Materials. In this scheme, coagulation can be expected to remove much of the high molecular weight dissolved organic material which has passed through the activated sludge process, such as refractory organic compounds and color causing compounds.

4. Heavy Metals. Each is a separate case, but many will be effectively removed by coagulation techniques.

5. N. B. If a settling tank or upflow clarifier is selected as the separation device, as is frequently done, a small concentration of otherwise coagulated impurities may wash over the weirs. The fact that they have been coagulated, however, renders them suitable for virtually complete removal in the granular filtration stage.

B. Scheme II: Physical-Chemical Treatment.

1. Suspended Solids. Similar to A1 above.

2. Phosphatic Pollutants. Similar to A2 above.

3. Organic Materials. Since in this scheme physical-chemical treatment has replaced secondary biological treatment appreciably greater concentrations of organic material will enter the coagulation stage. For a detailed discussion of their removal (with concurrent activated carbon adsorption) see W. J. Weber's report on adsorption in this series.

4. Heavy Metals. Similar to A4 above.

Polyvalent metal ions such as Al(III) and Fe(III) have proven themselves to be very effective in coagulation systems such as the one proposed in Scheme I. These two species are especially well suited for addition to biological wastewater treatment systems - - - particularly because the pH range of their greatest effectiveness (see section 2, Figures 6, 7, and 11) is close to the relatively neutral pH requirement of the nitrifying microorganisms. Concurrent use of metal ions with synthetic organic polyelectrolytes has also been shown to be effective (and economical) in some instances.

A number of chemical compounds are commercially available as sources of these metal ions. Alum ($Al_2(SO_4)_3 \cdot 17 H_2O$) and sodium aluminate ($NaAlO_3$) are currently the two most widely used sources of aluminum ion. Ferric chloride ($FeCl_3$) and ferric sulfate ($Fe_2(SO_4)_3$) are currently the two most widely used sources of iron. Since each of these chemical compounds has its own individual properties, there would be specific instances where one could be favored over the others (e. g., alum lowers pH, aluminate raises pH, etc.). In a general sense, however, they are basically quite similar and cost frequently becomes the deciding parameter. The cost per ton and cost per equivalent pound of metal ion for each of these compounds is shown in Table 2. The cost for synthetic organic polyelectrolytes will vary over a number of orders of magnitude. A very rough approximation is to increase the cost per pound of alum by a factor of ten to obtain a cost per pound of polyelectrolyte.

Table 2: Chemical and Cost Characteristics of Selected Inorganic Coagulants

Common Name	Chemical Formulation (Commercial)	Fraction Metal Ion	Cost Per Ton	Cost per lb. of Metal Ion (¢/lb metal ion)
Alum	$Al_2(SO_4)_3 \cdot 17H_2O$	0.083	$ 65	39.2
Sodium Aluminate	$NaAlO_3$	0.430	$150	17.4
Ferric Chloride	$FeCl_3$	0.345	$180	26.0
Ferric Sulfate	$Fe_2(SO_4)_3$	0.280	$ 50	8.9
Lime	CaO	0.714	$ 20	1.4

*Approximate local prices in Chicago area (June 1971); prices subject to appreciable variation particularly with respect to transportation distance.

Both lime and ferric chloride have been successfully employed in physical-chemical waste treatment systems such as the one proposed in Scheme II.

In terms of the Merricack Basin, the following coagulants are recommended.

- Scheme I. Preliminary recommendation is for a ferric salt due to generally better sludge settleability and compaction than with aluminum salt. In the absence of chemical recovery, ferric sulfate is more economical per pound of ferric ion, yet, with recovery ferric chloride is generally more suitable. Acid (base) adjustment may be desirable to minimize the requisite dosage of ferric sulfate.

- Scheme II. Preliminary recommendation is for lime. As noted

both lime and ferric chloride have been successfully employed. Lime appears to have distinct advantage here, however, not only because of cost, but also because of pH elevation which will offset subsequent pH reduction of break-point chlorination for ammonia nitrogen removal.

4. DESIGN CRITERIA

The parameters determining the requisite coagulant dosage for a given wastewater have been discussed in some detail in section 2 along with references to more detailed reports. Due to the variability and non-uniformity of wastewaters and the large number of parameters effecting the coagulation process, design of a coagulation process should be based on at least laboratory jar test data and preferably also on pilot plant data. In an attempt to give some idea as to what one might anticipate in the way of effective coagulants and their dosages in the Merrimack Basin, Table 3 below indicates the coagulants recommended for the three priority areas for the two proposed treatment schemes and their anticipated dose ranges.

Table 3 : Recommended Coagulants and Anticipated Dosages for the Treatment Schemes Proposed for the Three Priority Areas.

Priority Area	Scheme I Biological-Tertiary Treatment		Scheme II Physical-Chemical Treatment	
	Coagulant	Anticipated Dose Range	Coagulant	Anticipated Dose Range
Winnipesaukee Area	Ferric Sulfate	100-150 mg/l	Lime	200-250 mg/l
Lowell-Lawrence-Haverhill Area	Ferric Sulfate	150-200 mg/l	Lime	250-300 mg/l
Metropolitan Boston Area	Ferric Sulfate	125-175 mg/l	Lime	225-275 mg/l

The coagulants indicated in Table 3 were selected in terms of those that were believed to be most suited for the wastewater composition to be treated by the respective proposed treatment schemes. The anticipated coagulant concentration ranges were determined by approximating the amount of coagulant that would be required to achieve the effluent concentrations shown in Section 6, Tables 5 and 6 from the estimated influent composition shown in the same tables.

In the absence of jar test data (and/or pilot plant data) field operational experience has indicated that for the "conventional" three phase coagulation process (i.e., rapid mix, flocculation, and sedimentation) acceptable performance can normally be achieved with hydraulic detention times of 1-2 minutes for rapid mix, 15-30 minutes for flocculation, and 2-4 hours in a clarifier if quiescent type gravity sedimentation is selected for the separation device. The speed and extent of agitation (paddle stirring) to be provided in the flocculation basin can be estimated from a consideration of G factors (Fair, et al., 1968), and maintaining the G x t product in the range of 10^4 - 10^5. Surface overflow rates of 600-750 gallons/square foot/day are frequently used for the design of the clarification tanks with corresponding weir loadings of 15,000 gal./day/linear foot.

The controlling design parameters for Scheme I, with respect to coagulation, are associated principally with the design of the biological nitrification reactor and the corresponding sedimentation tank. In this regard it may be beneficial to present some operating data from a 50,000 gallon

per day EPA sponsored pilot plant study incorporating biological nitrification (Echelberger and Tenney, 1969, Guter, et al., 1971, Guter, 1971). Figure 12 shows a block diagram of the pilot plant flow scheme. The pilot plant is located at the City of South Bend Wastewater Treatment Plant. The wastewater being fed to the pilot plant receives the pretreatment of screening, grit removal, and comminution by the City's system prior to entering the pilot plant. As can be seen from Figure 12 the pilot plant basically consists of conventional activated sludge (with phosphorus removal by aluminum precipitation) followed by biological nitrification which is followed by biological denitrification (with methanol supplementation). Table 4 gives a listing of several wastewater constituents and their concentrations as flow progresses through each unit process of the pilot plant.

During the study period, the nitrification reactor had an average hydraulic detention time of 2.5 hours and the mixed liquor suspended solids concentration was maintained in the range 2000-2500 mg/l. The nitrification loading factor agreed closely with the previously reported value of 0.10 mg NH_4^+ N nitrified/day/mg MLVSS for this process. This value was observed at an average temperature of 18°C and pH = 7.5. It should be noted that reduction of either temperature or pH will appreciably decrease the efficiency of the process. A cell yield of approximately 0.05 mg of bio-mass/mg NH_4^+ N removed was observed in the system which also agrees closely with previously reported values. The amount of sludge recycled was established in order to maintain a solids retention time of approximately 5 days. The amount of oxygen required in

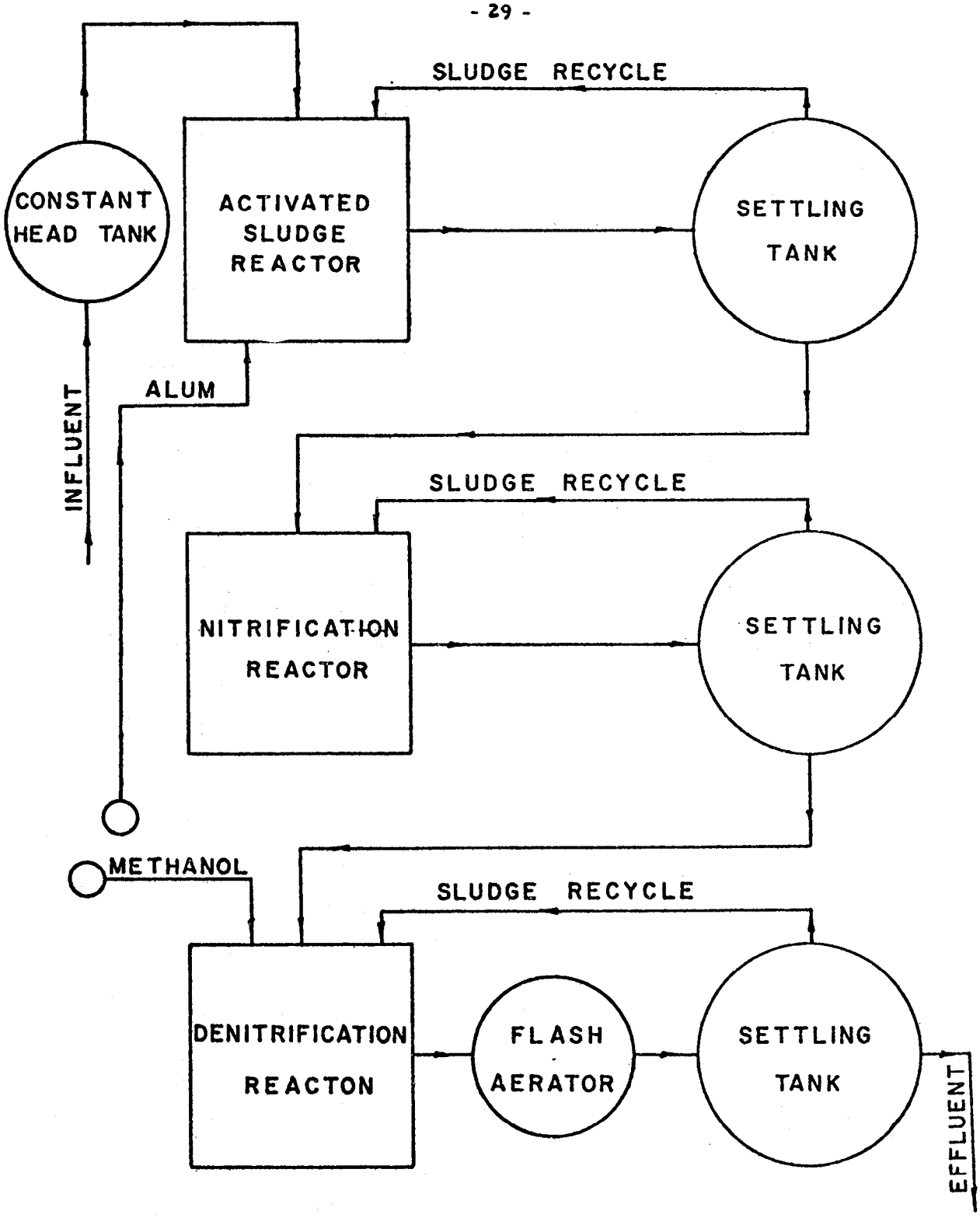


Figure 12: Block Diagram of Advanced Wastewater Treatment Pilot Plant.

TABLE 4: SUMMARY OF NOTRE DAME PILOT PLANT OPERATION, MAY, 1970

Wastewater Parameter	Raw Wastewater	Activated Sludge	Nitrified Effluent	Denitrified Effluent
COD				
Average	340	85	54	55
Range	180-565	38-107	29-95	36-97
Ammonia-N				
Average	6.0	4.3	0.3	0.3
Range	1.6-10.2	1.9-7.1	0.1-1.2	0.0-1.5
Nitrite and Nitrate-N				
Average	0.8	0.8	5.7	0.0
Range	0.3-1.4	0.3-1.3	3.4-10.0	0.0-0.0
Ortho-P				
Average	2.2	0.4	0.5	0.4
Range	0.6-3.5	0.1-1.1	0.3-1.0	0.3-0.7
Total-P				
Average	3.6	0.8	1.1	1.0
Range	3.1-5.9	0.2-2.1	0.5-3.0	0.7-2.1
Suspended Solids				
Average	159	36	28	22
Range	74-326	20-57	8-71	10-42

Note: All concentration values expressed in milligrams per liter.

the reactor is determined primarily by the influent ammonia concentration - - - 1 mg/l of NH_4^+ as nitrogen requiring approximately 4.6 mg/l of oxygen.

The sludge was observed to settle well and possessed a sludge volume index in the range of 50-60. The clarifier was designed as a circular rim flow upflow type with an average overflow rate of approximately 575 gallons/square foot/day.

5. PRE-TREATMENTS REQUIRED

In Scheme I (Biological-Tertiary Treatment) it is assumed that the influent to the biological nitrification-coagulation reactor is of equivalent quality as the effluent from a conventional secondary wastewater treatment plant (refer to Section 1 - Figure 1). Thus such processes as screening, grit removal, comminution, primary sedimentation with scum removal, and removal of dissolved organic impurities (i. e., activated sludge or trickling filters) would proceed the biological nitrification-coagulation reactor. No specialized pre-treatment as such would be required (N. B., it is presumed that discharges to the sewers will be controlled through Base Level Treatment ordinances, etc. in order to preclude shock or toxic loadings). Of course, it would be desirable to have all the primary and secondary pre-treatment operations cited above, but operationally the flexibility of the coagulation process is such that it can function effectively with minor adjustments in the absence of many of these "pre-treatments." It is my opinion (although there is some contradicting information in the literature) that the one pre-treatment process absolutely necessary to this scheme is the removal of the dissolved organic pollutants, otherwise biological nitrification can not proceed effectively.

For Scheme II (Physical-Chemical Treatment) it is most desirable (although again not imperative) that the influent to the coagulation process be the equivalent in quality as the effluent from a conventional primary wastewater treatment plant. Thus such processes as screening, grit re-

moval, comminution and primary sedimentation with scum removal would be desirable to have preceeding the coagulation process.

6. EFFLUENT QUALITIES

The effluent qualities anticipated from the two schemes proposed are shown for each of the three priority areas in Tables 5 and 6 . The anticipated influent concentrations to the two systems for each of the three priority areas are also shown.

It will be observed from a comparison of these two tables that the anticipated effluent quality from the coagulation process in each of the two schemes suggested is expected to be quite similar for the parameters shown. The major differences are:

A. Scheme I: Biological-Tertiary Treatment.

1. Nitrogen will almost entirely be in the nitrate form allowing essentially complete nitrogen removal to occur by denitrification on the carbon beds.
2. Two stage operation with individual settling tanks are required to provide optimum environmental conditions for the respective microbial species.

B. Scheme II: Physical-Chemical Treatment

1. Biological treatment is eliminated.
2. Nitrogen removal is accomplished by chlorination - - - only ammonia nitrogen forms can be removed in this fashion, however.

Table 5 : Anticipated Influent and Effluent Concentrations to a Coagulation (Nitrification) Process Following Activated Sludge Treatment (Scheme I).

Parameter	Boston		Lowell-Lawrence-Haverhill		Winnipesaukee	
	Coagulation Influent	Coagulation Effluent	Coagulation Influent	Coagulation Effluent	Coagulation Influent	Coagulation Effluent
BOD_5 - mg/l	25	15	25	17	25	17
Susp. Solids - mg/l	25	20	25	20	25	20
COD - mg/l	70	60	140	120	75	60
Total Nitrogen mg/l as N	20	18	10	10	25	23
ortho Phosphate mg/l as P	10	1	8	1	10	1
Total Dissolved Solids - mg/l	275	275	320	320	275	275

Table 6 : Anticipated Influent and Effluent Concentrations to a Coagulation Process in a Physical Chemical Treatment Process (Scheme II).

Parameter	Boston		Lowell-Lawrence-Haverhill		Winnipesaukee	
	Coagulation Influent	Coagulation Effluent	Coagulation Influent	Coagulation Effluent	Coagulation Influent	Coagulation Effluent
BOD_5 - mg/l	175	30	215	75	164	30
Susp. Solids - mg/l	260	15	245	20	158	15
COD - mg/l	260	70	320	130	275	70
Total Nitrogen mg/l as N	25	20	15	12	30	25
ortho Phosphate mg/l as P	12	1	10	1	15	1
Total Dissolved Solids - mg/l	300	350	335	385	300	350

7. EXPERIENCE ON PROCESS TO DATE

The following references represent typical references related to the subject of this report and each was discussed in the body of this report. Numerous other references exist, many of which are included in the bibliographies of the references cited here:

American Water Works Association Coagulation Committee (J. E. Singley, Chairman), "State of the Art of Coagulation - - Mechanisms and Stoichiometry." Journal American Water Works Association, 63, 99-108 (1971).

Barth, E. F., Brenner, R. C., and Lewis, R. F., "Chemical-Biological Control of Nitrogen and Phosphorus in Wastewater Effluent." Journal Water Pollution Control Federation, 40, 2040-2054 (1968).

Barth, E. F., Jackson, B. N., Lewis, R. F., and Brenner, R. C., "Phosphorus Removal from Wastewater by Direct Dosing of Aluminate to a Trickling Filter." Journal Water Pollution Control Federation, 41, 1932-1942 (1969).

Echelberger, W. F., Jr. and Tenney, M. W., "Wastewater Treatment for Complete Nutrient Removal." Water and Sewage Works, 116, 396-402 (1969).

Environmental Protection Agency, "Advanced Waste Treatment and Water Reuse Symposium" Volumes I and II (1971).

Fair, G. M., Geyer, J. C., and Okun, D. A., "Water and Wastewater Engineering." Volume II. John Wiley and Sons, Inc., New York (1968).

Guter, K. J., Echelberger, W. F., Jr., and Tenney, M. W., "Evaluation of Anaerobic Denitirification Processes - Discussion." Journal Sanitary Engineering Division American Society of Civil Engineers., 97, 109-111 (1971).

Guter, K. J., "Removal of Organic and Inorganic Pollutants from Wastewater by Combined Chemical and Biological Treatment." unpublished Ph. D. thesis, University of Notre Dame (1971).

Hall, M. W. and Engelbrecht, R. S., "Phosphorus Removal - Past, Present, and Future." Water and Wastes Engineering, 6, 50-53 (August 1969).

Laughlin, J. E., "Designing for Phosphorus Removal with Metal Salts and Polymers." paper presented as part of the EPA Design Seminar for Wastewater Treatment Facilities Charlottesville, Virginia (June 1971).

Little, L. W., "Phosphorus in Water and Wastewater - An Annotated Selected Bibliography." University of North Carolina Research Center Report No. 11, Chapel Hill, North Carolina (November 1970).

McCarty, P. L., Beck. L., and St. Amant, P., "Biological Denitrification of Wastewaters by Addition of Organic Materials." 24th Purdue Industrial Waste Conference, 1271-1285 (1969).

Nesbitt, J. B., "Phosphorus Removal - The State of the Art." Journal Water Pollution Control Federation, 41, 701-713 (1969).

O'Melia, C. R., "A Review of the Coagulation Process." Public Works, 100, 87- (1969).

Slechta, A. F. and Culp, G. L., "Water Reclamation Studies at the South Tahoe Public Utility District." Journal Water Pollution Control Federation, 39, 787-814 (1967).

Smith, R., "Cost of Conventional and Advanced Treatment of Wastewater." Journal Water Pollution Control Federation, 40, 1546-1574 (1968).

Tenney, M. W. and Stumm, W., "Chemical Flocculation of Microorganisms in Biological Waste Treatment." Journal Water Pollution Control Federation, 37, 1370-1388 (1965).

Theis, T. L., Singer, P. C., Echelberger, W. F., Jr. and Tenney, M. W., "Phosphate Removal: Summary of Papers - Discussion." Journal Sanitary Engineering Division, American Society of Civil Engineers, 96, 1004-1009 (1970).

Weber, W. J., Jr., Hopkins, C. B., and Bloom, R., Jr., "Physiochemical Treatment of Wastewater." Journal Water Pollution Control Federation, 42, 83-99 (1970).

Wild, H. E., Sawyer, C. N., McMahon, T. C., "Factors Affecting Nitrification Kinetics" paper presented at the 43rd Annual Conference of the Water Pollution Control Federation, Boston, Massachusetts (October, 1970).

Zuckerman, M M. and Molof, A. H., "High Quality Reuse Water by Chemical-Physical Wastewater Treatment." Journal Water Pollution Control Federation, 42, 437-456 (1970).

8. CAPITAL AND OPERATING COSTS

Due to the number of alternatives that exist for separate or combined treatment in or between the areas of concern in the Merrimack River Basin, approximate cost figures for the two treatment schemes considered are given at various flow rates. Extimated cost figures have been prepared for 1 MGD, 10 MGD, 100 MGD, and 1,000 MGD plants for the two proposed treatment schemes and are presented in Tables 7 and 8, respectively and plotted in a fashion similar to Smith's (Smith, 1968) in Figures 13 and 14, respectively.

As can be observed from these estimates, a major operating expense of the coagulation process is, of course, associated with the purchase of the chemical coagulants. Thus, any modification in coagulant dose, coagulant recovery, etc. will have a marked effect on the cost per 1000 gallons of water treated. Recovery and recycle of the chemical coagulants would appreciably reduce the cost per 1000 gallons of water treated.

Table 7: Estimated Capital and Treatment Cost of Coagulation (with Clarification) for Selected Treatment Plant Capacities in Scheme I (Biological-Tertiary)

Cost Elements	1 MGD		10 MGD		100 MGD		1000 MGD	
	Cost/yr	¢/kgal	Cost/yr	¢/kgal	Cost/yr	¢/kgal	Cost/yr	¢/kgal
Capital Costs								
1) Plant & Equipment	9,200	2.5	64,000	1.75	440,000	1.2	3,200,000	0.85
a) Tanks								
b) Aeration Equipment								
c) Sludge Collection								
d) Chemical Feeders					440,000	1.2	3,200,000	0.85
Total Capital Cost	9,200	2.5	64,000	1.75				
Operation & Maintenance Costs								
2) Operating & Maintenance Labor	13,000	3.5	90,000	2.5	610,000	1.7	3,250,000	0.9
3) General & Administrative Overhead	4,000	1.1	20,000	0.5	90,000	0.3	750,000	0.2
4) Chemicals	11,300	3.1	113,000	3.1	1,131,500	3.1	11,315,000	3.1
5) Electric Power (0.01¢/KWH)	3,600	1.0	36,000	1.0	360,000	1.0	3,600,000	1.0
Total Operating & Maintenance Costs	31,900	8.7	259,000	7.1	2,190,000	6.1	18,900,000	5.20
Total Water Costs	41,100	11.2	323,000	8.85	2,630,000	7.3	22,100,000	6.05

Table 8: Estimated Capital and Treatment Costs of Coagulation (with Clarification) for Selected Treatment Plant Capacities in Scheme II (Physical - Chemical)

Cost Elements	1 MGD		10 MGD		100 MGD		1000 MGD	
	Cost/yr	¢/kgal	Cost/yr	¢/kgal	Cost/yr	¢/kgal	Cost/yr	¢/kgal
Capital Costs								
1) Plant & Equipment	4,160	1.3	31,200	1.1	256,000	0.7	2,160,000	0.5
a) Rapid Mixers								
b) Flocculation								
c) Sludge Collection								
d) Tanks								
e) Chemical Feeders								
Total Capital Cost	4,160	1.3	31,200	1.1	256,000	0.7	2,160,000	0.5
Operation & Maintenance Costs								
2) Operating & Maintenance Labor	9,000	2.5	24,000	0.65	132,000	0.4	1,500,000	0.4
3) General & Administrative Overhead	2,500	0.7	8,000	0.2	40,000	0.1	350,000	0.02
4) Chemicals	7,300	2.10	73,000	2.10	730,000	2.10	7,300,000	2.10
5) Electric Power (0.01¢/KWH)	400	.01	2,000	.05	30,000	.08	330,000	.09
Total Operating & Maintenance Costs	19,200	5.31	107,000	3.00	932,000	2.68	9,480,000	2.61
Total Water Costs	23,360	6.61	138,200	4.1	1,188,000	3.38	11,640,000	3.11

9. SAMPLE CALCULATIONS

A. See Figure # 14 Total Plant Cost = $52,000

B. Amortize at 7% for 30 years

$$R = P\left[\frac{i(1+i)^n}{(1+i)^n - 1}\right]$$

$$R = 52,000\left[\frac{0.07(1.07)^{30}}{(1.07)^{30} - 1}\right] = 4,160/\text{yr}$$

$$R = 4,160 \times \frac{1}{365} \times \frac{1,000}{1,000,000} = 1.3¢/1000 \text{ gal}$$

C. Chemical Cost

1) Lime (250 mg/l as CaO)

1 mg/l = 8.34 lb/MG

$$250 \times 8.34 \text{ lb} \times 1 \text{ MG} \times \frac{1 \text{ ton}}{2000} = 1.042 \text{ ton}$$

Cost = $20/ton x 1.042 = 20.84

$$= 20.84 \times \frac{1000}{1,000,000} = \$0.021/1000 \text{ gal} = 2.1¢/1000 \text{ gal}$$

2) Ferric Sulfate (150 mg/l)

1 mg/l = 8.34 lb/MG

$$150 \times 8.34 \text{ lb} \times 1 \text{ MG} \times \frac{1 \text{ ton}}{2000 \text{ lb}} = 0.625 \text{ tons}$$

Cost = $50/ton x 0.625 = 31.25/MG

$$= 31.25 \times \frac{1,000}{1,000,000} = \$0.031/1000 \text{ gal} = 3.1¢/1000 \text{ gal}$$

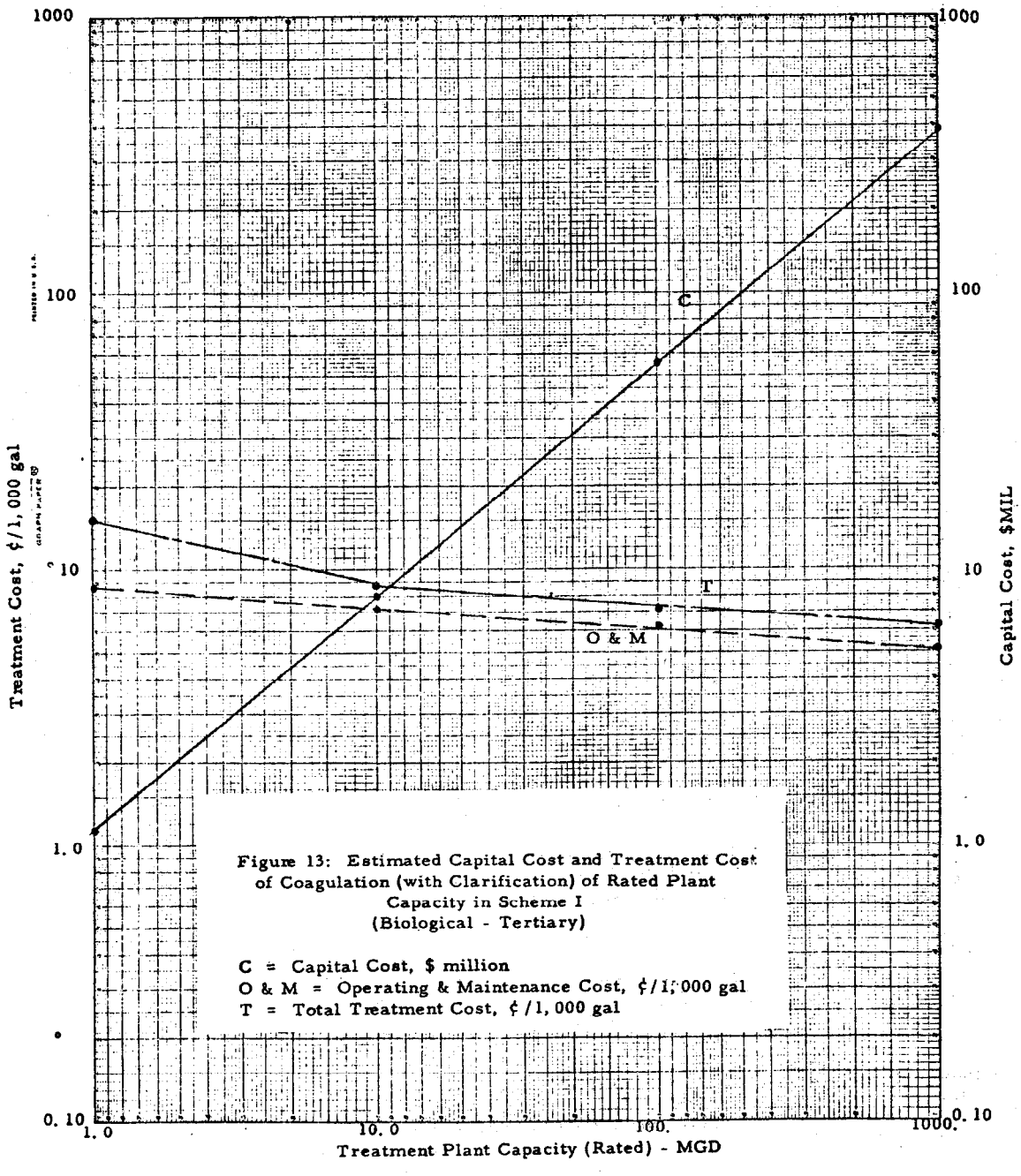


Figure 13: Estimated Capital Cost and Treatment Cost of Coagulation (with Clarification) of Rated Plant Capacity in Scheme I (Biological - Tertiary)

C = Capital Cost, $ million
O & M = Operating & Maintenance Cost, ¢/1,000 gal
T = Total Treatment Cost, ¢/1,000 gal

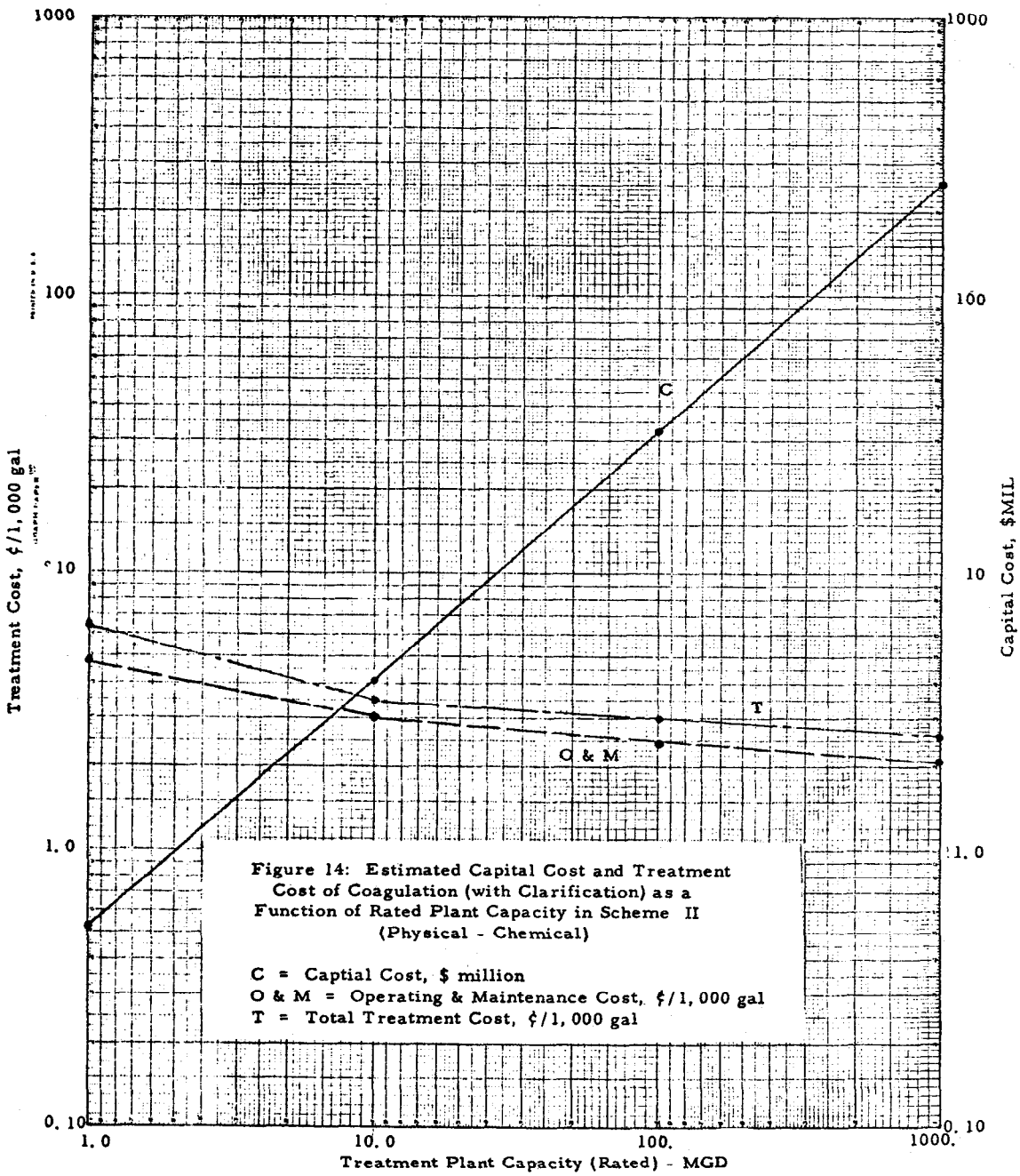


Figure 14: Estimated Capital Cost and Treatment Cost of Coagulation (with Clarification) as a Function of Rated Plant Capacity in Scheme II (Physical - Chemical)

C = Captial Cost, $ million
O & M = Operating & Maintenance Cost, ¢/1,000 gal
T = Total Treatment Cost, ¢/1,000 gal

10. SCALE UP FACTORS

Scale up of coagulation systems is virtually direct. Experience has indicated that laboratory scale and pilot plant data can be scaled up almost directly to field installations of any given size. For scale up considerations, however, individual units in the range of 1 - 10 MGD are generally the maximum size designed. Proper design of wastewater treatment plant units necessitates parallel operation of similar units (split flow) for flexibility in smaller plants (i. e., <10 MGD) and for efficient operation and management in the larger plants. Thus, scale up considerations would be applicable to individual units up to and including approximately 10 MGD total capacity.

Scale up is straightforward in coagulation primarily because of the fundementals of the process. As described in detail in earlier sections, laboratory (and pilot plant) studies should first determine those coagulants that will work effectively for the particular wastewater under consideration and then determine the influence of such parameters as pH, alkalinity, impurity concentration, agitation, temperature, etc. on the requisite coagulant dose. Ultimately the laboratory (and pilot plant) data would result in specifying operating ranges so that such items as tank volumes, paddle stirrers, chemical feeders, etc. could be sized. Any small discrepencies that might result from scale-up can be completely compensated for in the field by adjusting such parameters as the chemical feeders, the extent of agitation, etc.

11. PROCESS DEPENDABILITY

The coagulation process as such is a most dependable process. The fact that it is basically a chemical treatment process and that the operation is confined to a given tank or reactor volume provides for exceptional control. Excellent instrumentation is available for instaneous monitoring of both influent and effluent properties as well as in situ monitoring of the fundemental properties (or underlying mechanisms) occuring in the reactor. Variations in the influent composition will not shock this system as the chemical(s) can be adjusted accordingly. Appreciably large increases in flow can impair the process by reducing hydraulic detention times in the reactors and increasing the overflow rates in sedimentation tanks. Sudden lowering of temperatures can also impair the functioning of the process; proper below ground positioning of the tanks normally eliminates this difficulty, however.

12. MATRIX SUMMARY

A matrix type summary is given below in Table 9 for the coagulation process as it applies to both proposed schemes. It will be observed from this table that the process is rated extremely high and is inherently well suited for utilization in both treatment schemes. There are virtually no difficulties anticipated with the process in terms of what it is designed to remove from the wastewaters. As indicated the scale up is virtually direct from laboratory through pilot plant through 1 MGD beyond 100 MGD. The dependability over time, flexibility, tolerance to influent variability, and effluent variability are all rated exceptional primarily because coagulation is a chemical treatment process which is inherently flexible and thus is capable of adjusting to variations in influent composition and will continously produce an effluent of excellent quality. Virtually no ancillary processes are required except that the influent to the biological nitrification - coagulation process in Scheme I must be low in organic carbon. The process costs, efficiency, and operation difficulties are similarily rated extremely favorable.

Table 9: MATRIX SUMMARY FOR COAGULATION*

Parameter	Target Year		
	1975	1990	2020
Scale-up	Direct	Direct	Direct
Dependability over Time	1	1	1
Flexibility (in relation to other processes)	1	1	1
Tolerance to Influent Variability	1	1	1
Effluent Variability	1	1	1
Ancillary Processes Required	virtually none	virtually none	virtually none
Costs	1	1	1
Process Efficiency	1	1	1
Operational Difficulty	1	1	1

*Rated on a scale of 1 (best) to 5 (worst)

Preliminary Examination

INSTREAM AERATION ALTERNATIVES ON THE MERRIMACK RIVER

William Whipple Jr.
Water Resources Research Institute

Shaw L. Yu
Department of Civil and Environmental Engineering

July 9, 1971

Water Resources Research Institute
Rutgers University
New Brunswick, N. J. 08903

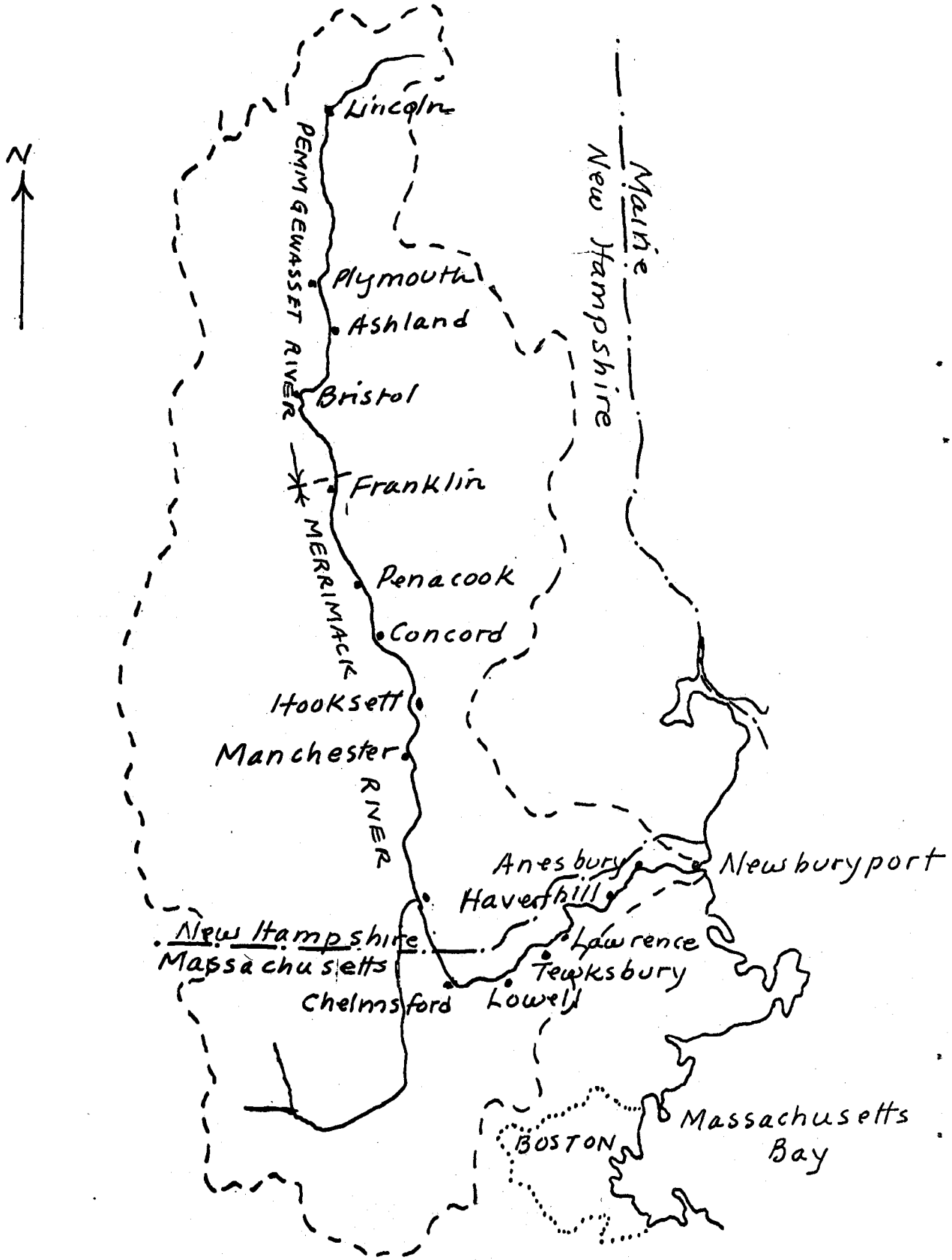


FIG. 1

THE MERRIMACK RIVER BASIN

1. Introduction

This is a preliminary study of the possibilities of improving dissolved oxygen levels on the Merrimack River, New Hampshire and Massachusetts, by means of instream aeration, employed as a supplement to a program of secondary waste treatment. It was undertaken by the Water Resources Research Institute, Rutgers University, for the Corps of Engineers, New England Division, Waltham, Massachusetts. The study was conducted by Wm. Whipple, Jr. and S. L. Yu, during the period 9 June 1971 to 9 July 1971. Required data as to pollution loadings and other characteristics of the river were provided by the Corps of Engineers, and included work previously done by others, which acknowledged where practicable.

2. The Merrimack River

The basin extends from the White Mountains in New Hampshire southerly into Massachusetts. It then turns eastward and flows 45 miles to the sea, the lower 22 miles being tidal. The drainage area is 5010 square miles, and the 1960 population within the basin was estimated to be 1,072,000, of which 727,000 were in Massachusetts. The three largest towns are Manchester, N.H. [88,282], Lowell, Mass. [92,107] and Lawrence, Mass. [70,933]. See Map, Fig. 1. Most of the towns and industries in the basin discharge their wastes into the rivers without any treatment whatsoever.

3. Information Available

Information provided as basis for the study included the following data grouped as indicated below.

*A. Flow probability curves for the river at various points.

B. River profiles, flood frequency, times of travel for various

---

*Referred to as Attachment A, B, etc., although not attached to this report.

reaches of the river, in each case related to river discharge.

C. Stream parameters for individual reaches for the 7-day minimum flow. These included the usual parameters relevant to biochemical action, exclusive of that for nitrification.

D. Estimated pollution loads, cities, towns and industries including both primary and nitrogenous BOD. Estimates of future served population. Identification of reaches.

E. Population and sewage flow trends for various towns and cities.

4. Mathematical Modeling of the Dissolved Oxygen Profile

Based on the information provided by the Corps of Engineers, it was learned that the Merrimack River had been modeled by using Camp's approach (1), and low flows of 10% probability of occurrence. The resulted DO profile for the entire river is shown in Figure 2. Also, the Federal Water Pollution Control Administration (2) has studied the lower Merrimack using Camp's equation. The results of the above two studies provide a basis for verifying the mathematical model used in the present study. However, not all of the assumptions and data used in these prior studies were available.

In view of the fact that a substantial amount of nitrogenous biochemical oxygen demand loadings is discharged into the Merrimack and that the travel time from the origin to the mouth exceeds thirty days, it was felt that the effect of nitrification should be included in the model. It appears that the use of the modified DO equation by O'Connor (3) would be appropriate. The equation is as follows:

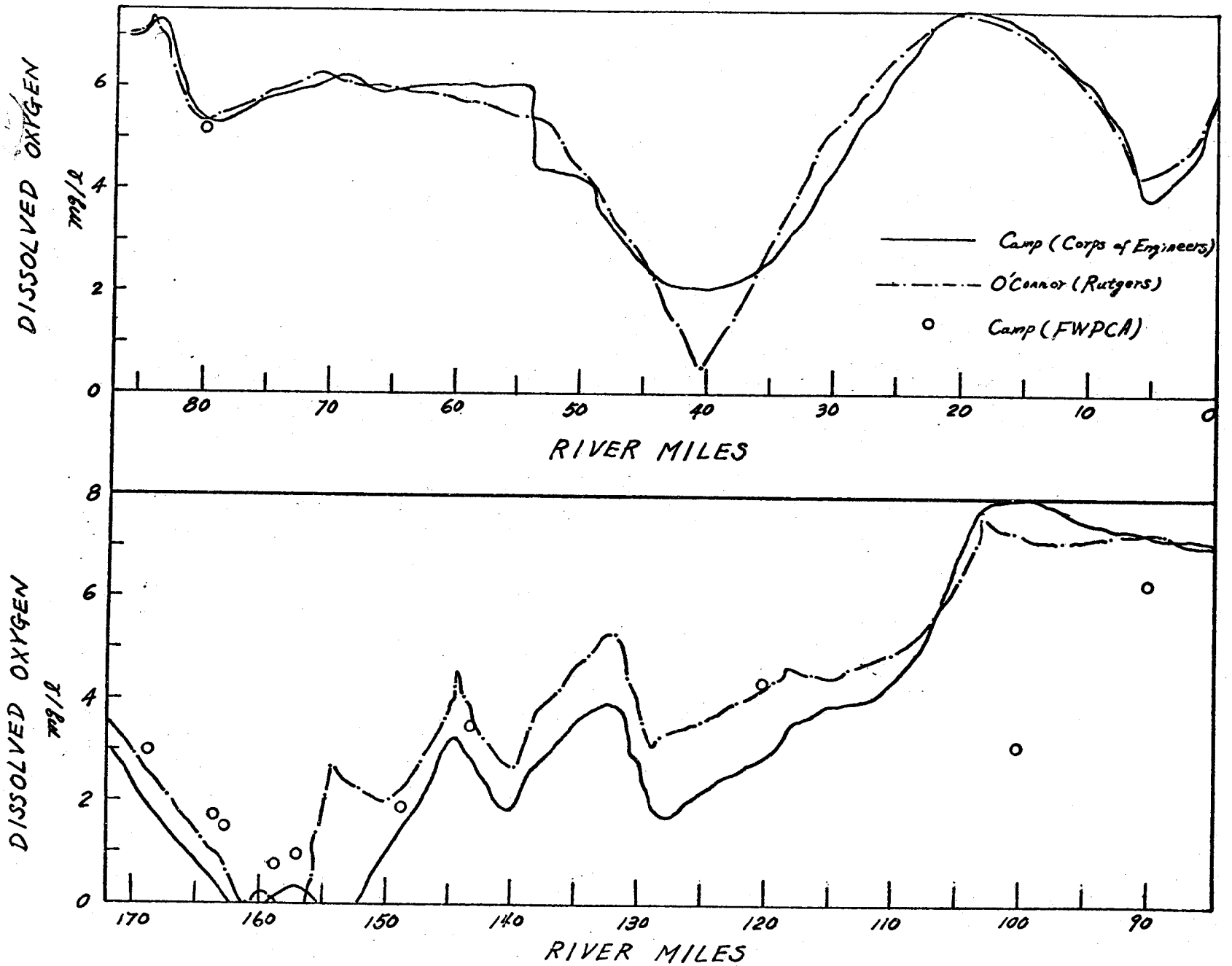


FIG. 2 MODEL VARIFICATION

$$D_t = \frac{K_d L_o}{K_a - K_r}(e^{-K_r t} - e^{-K_a t}) + \frac{K_n N_o}{K_a - K_n}(e^{-K_n t} - e^{-K_a t})$$

$$+ D_o(e^{-K_a t}) + \frac{S}{K_a}(1 - e^{-K_a t}) - \frac{\overline{P-R}}{K_a}(1 - e^{-K_a t}) \qquad [1]$$

in which

D_t = the oxygen deficit at time t downstream from point of waste discharge

K_d = the deoxygenation constant

L_o = the initial ultimate BOD

K_a = the atmospheric reaeration constant

K_r = the BOD removal constant

N_o = the initial ultimate nitrogenous oxygen demand

K_n = the nitrogenous oxidation constant

D_o = the initial oxygen deficit

S = the benthol oxygen demand

$\overline{P-R}$ = the net production of photosynthetic oxygen

5. Parameter Estimation

Most of the parameters needed for using Equation [1] were given in Attachment C, provided by the Corps of Engineers. However, the following assumptions were necessary:

(1) $K_n = K_1$; the same assumption has been employed elsewhere.(4)

(2) $K_r = K_d = K_1 + K_3$

6. Model Verification

The waste loadings for the year 1970 listed in Attachment D were used for verifying the model. At the request of the Corps of Engineers, it was assumed that the river had an average temperature

of 26°C, which gave a saturation oxygen value of 8.0 mg/ℓ, although the profiles given indicate a saturation value of water for oxygen of at least 9.0 mg/ℓ. (5)

It is well known that there is uaually a time lag after introduction of ammonia into a stream before the nitrification process takes place. In view of the fact that most of the parameters in Equation [1] are already given, it was decided to vary the lag time in order to best reproduce the DO profile shown in Fig. 2.

For modelling purposes, the river is divided into 51 reaches. The DO values for each reach were computed by Equation [1] on the Rutgers IBM 360/67 computer. The nitrification term was treated by putting an equivalent imaginary load at a point downstream with travel time equal to the lag time. Several values of lag time, namely, 10, 7, and 5 days, were tried.

It was found that the DO profile with a lag time of 5 days for nitrification resemble most closely the original profile. See Fig.2. DO observations made by the FWPCA during the summer of 1964-65 are also plotted on Fig. 2 for comparison and show a reasonably close correspondence.

7. Instream Aeration Alternative

With existing level of waste treatment, the DO profile for 1970 waste loadings is shown in Figure 3. This profile was prepared by using the O'Connor's model with a lag time of 5 days for nitrification. Also since it has been shown (6) that a substantial amount of unrecorded waste loadings are apt to enter a river, particularly after rainfall, these results are based upon an assumption of a 25% unrecorded loading at each point source. The data given for Nashua included the flow of the river, and therefore, included unrecorded BOD already.

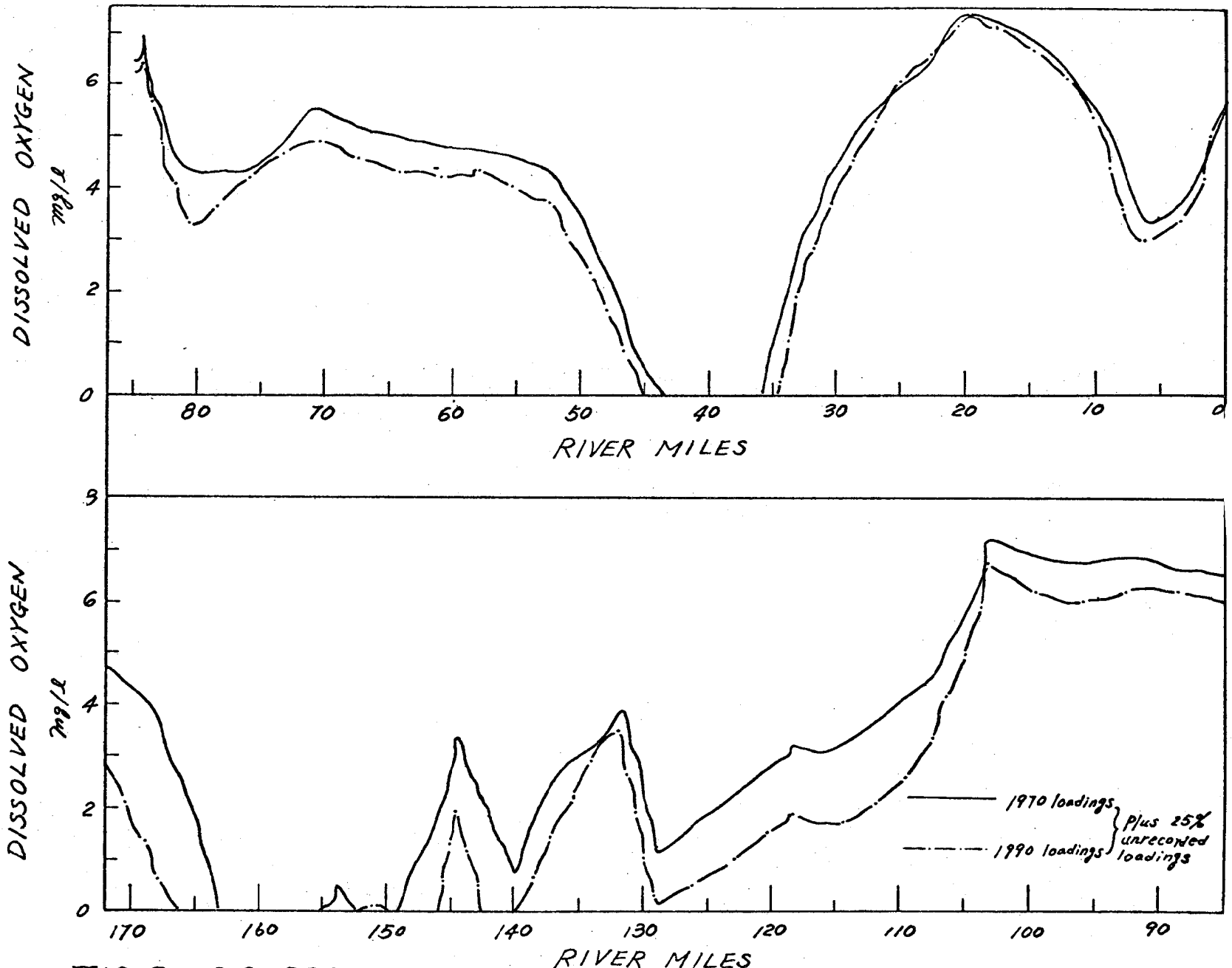


FIG. 3 DO PROFILES WITH EXISTING TREATMENT

To estimate the amount of instream aeration capacity required, it s assumed that a uniform basin-wide secondary waste treatment system is provided. A BOD removal efficiency of 80% was assumed. As for nitrification, an efficiency of between 30 and 40 per cent may be practicable (7). Therefore, both values were used in the computation.

The DO standard requires 5 mg/ℓ for some parts of the river and 6 mg/ℓ for others, as given by the Corps. Also, a uniform 5 mg/ℓ standard was used for comparison. The aerator considered is the electric-drive surface mechanical aerator with 75 shaft horsepower. Estimates are based upon the more expensive slow-moving type.

The number of aerators required for various conditions is given in Table 1. Figure 3 shows low DO conditions, with existing degree of treatment, and loadings of 1970 and of 1990, with 25% unrecorded loading added. Figs. 4 and 4A show the DO profiles for 1970 and 1990 that would be resulted from a general secondary treatment averaging 80% removal of primary BOD and 30% removal of secondary BOD, without in-stream aeration. Figure 5* shows the DO profiles for a Case 1 remedial program (Table 1). This consists of general secondary treatment averaging 80% removal of primary BOD and 40% removal of secondary BOD, and additional aerators, 3 in 1970 and 12 in 1990, as required to meet a uniform standard of 5.0 mg/ℓ. Figure 6* shows the same conditions again with a Case 1 remedial program, but in this instance designed to meet current water quality standards of 6.0 mg/ℓ in certain parts of the stream. The number of aerators required is increased to 8 for 1970 and 25 for 1990. The requirements of Cases 2 and 3, as shown in Table 1, are not illustrated.

8. Cost Estimates

Provisions for by-passing small boat navigations would be

* The profiles in Figs. 5 and 6 represent conditions with 25% unrecorded loading added.

TABLE 1

ESTIMATES OF THE NUMBER OF AERATORS (75-hp) REQUIRED TO MAINTAIN THE DO STANDARDS ($C_{min.}$) SPECIFIED

Case No.	1970		1990	
	$C_{min.}$ = 5 mg/ℓ	$C_{min.}$ = 5 or 6 mg/ℓ	$C_{min.}$ = 5 mg/ℓ	$C_{min.}$ = 5 or 6 mg/ℓ
1. Secondary treatment (80% bod, 40% nod) removal, plus 25% unrecorded loadings	3	8	12	25
2. Secondary treatment (80% bod, 30% nod) removal, plus 25% unrecorded loadings	6	15	18	35
3. Secondary treatment (80% bod, 30% nod) removal, without unrecorded loadings	1	5	9	20

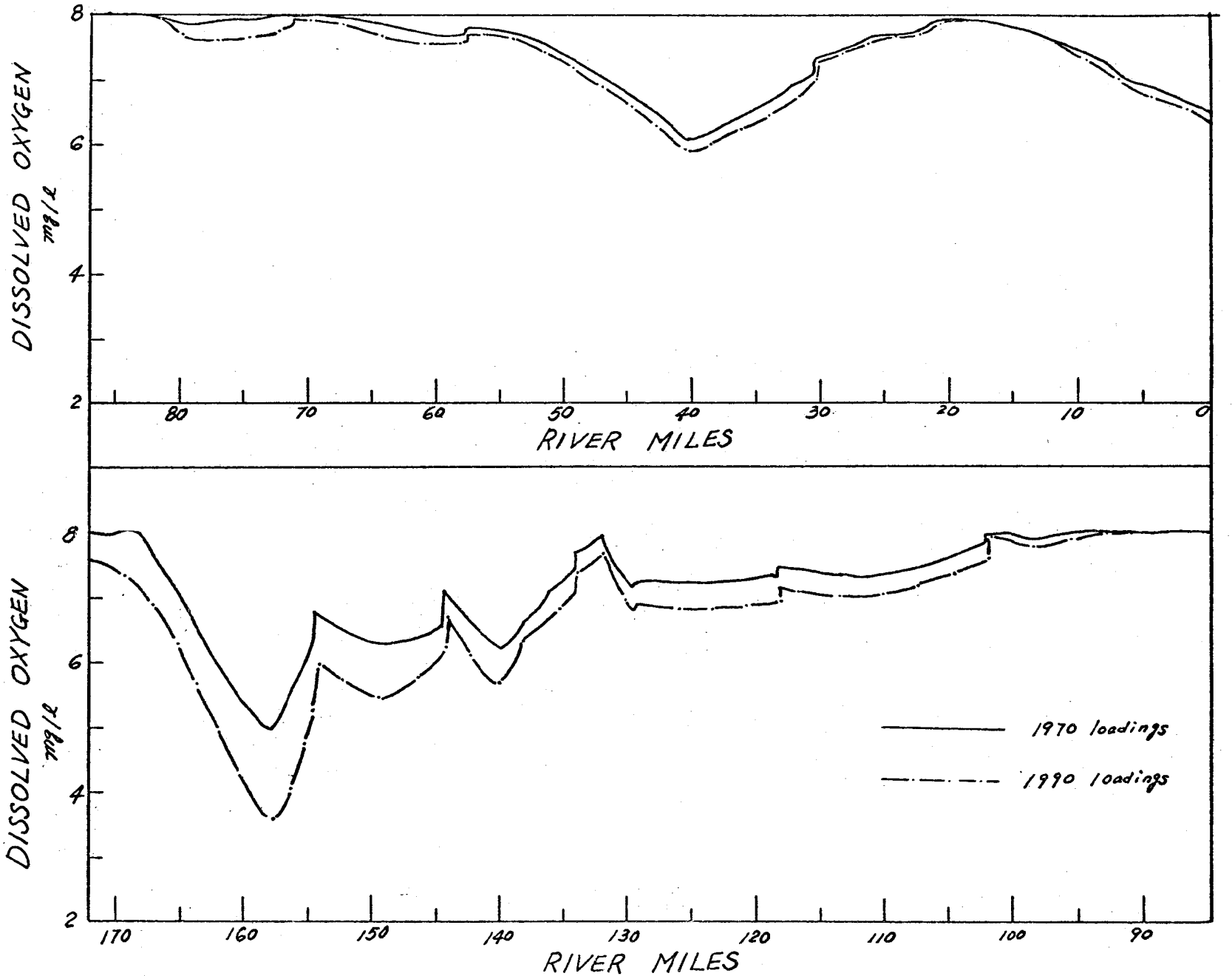


FIG. 4 DO PROFILES WITH SECONDARY TREATMENT

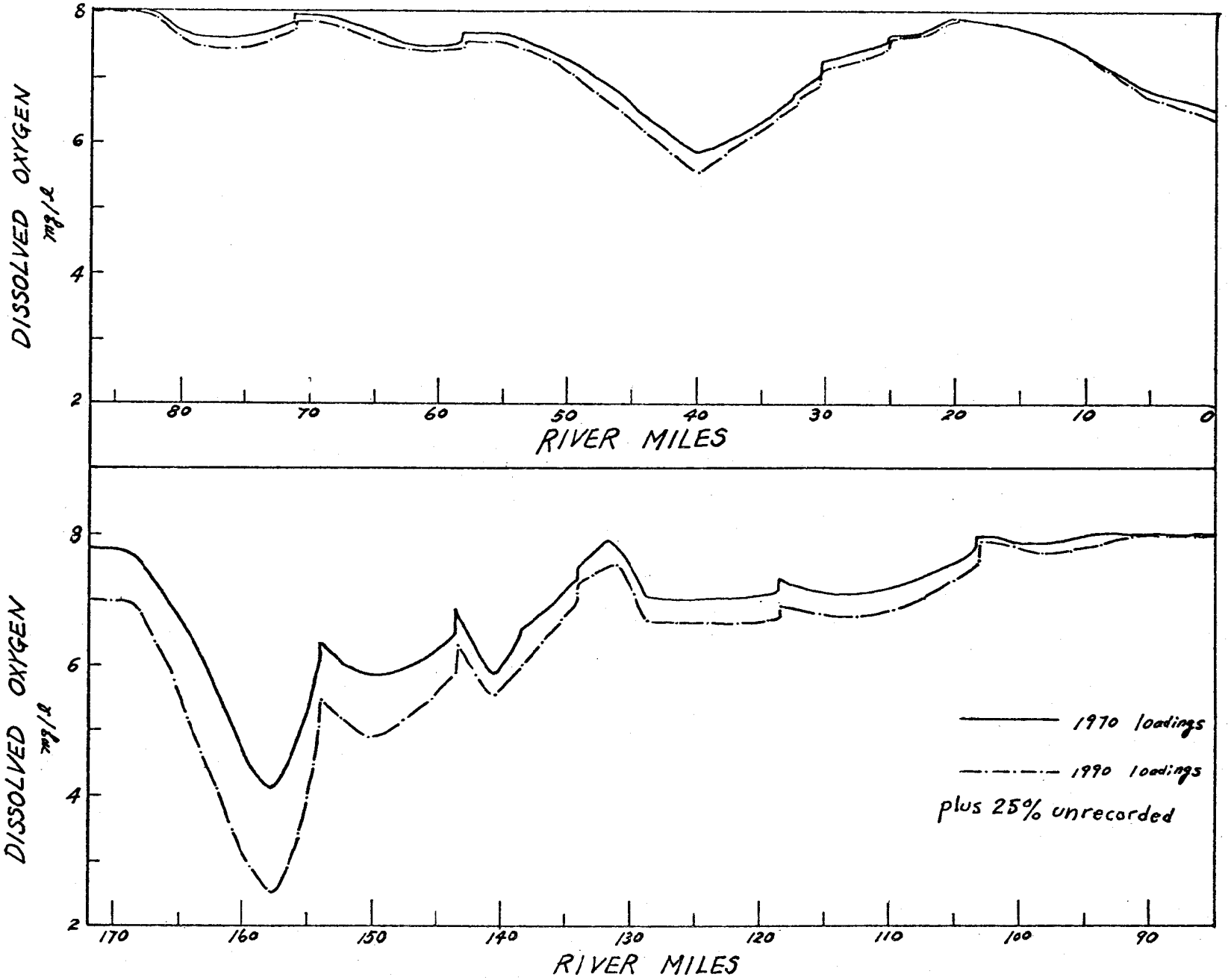


FIG. 4A DO PROFILES WITH SECONDARY TREATMENT PLUS 25% UNRECODED LOADING

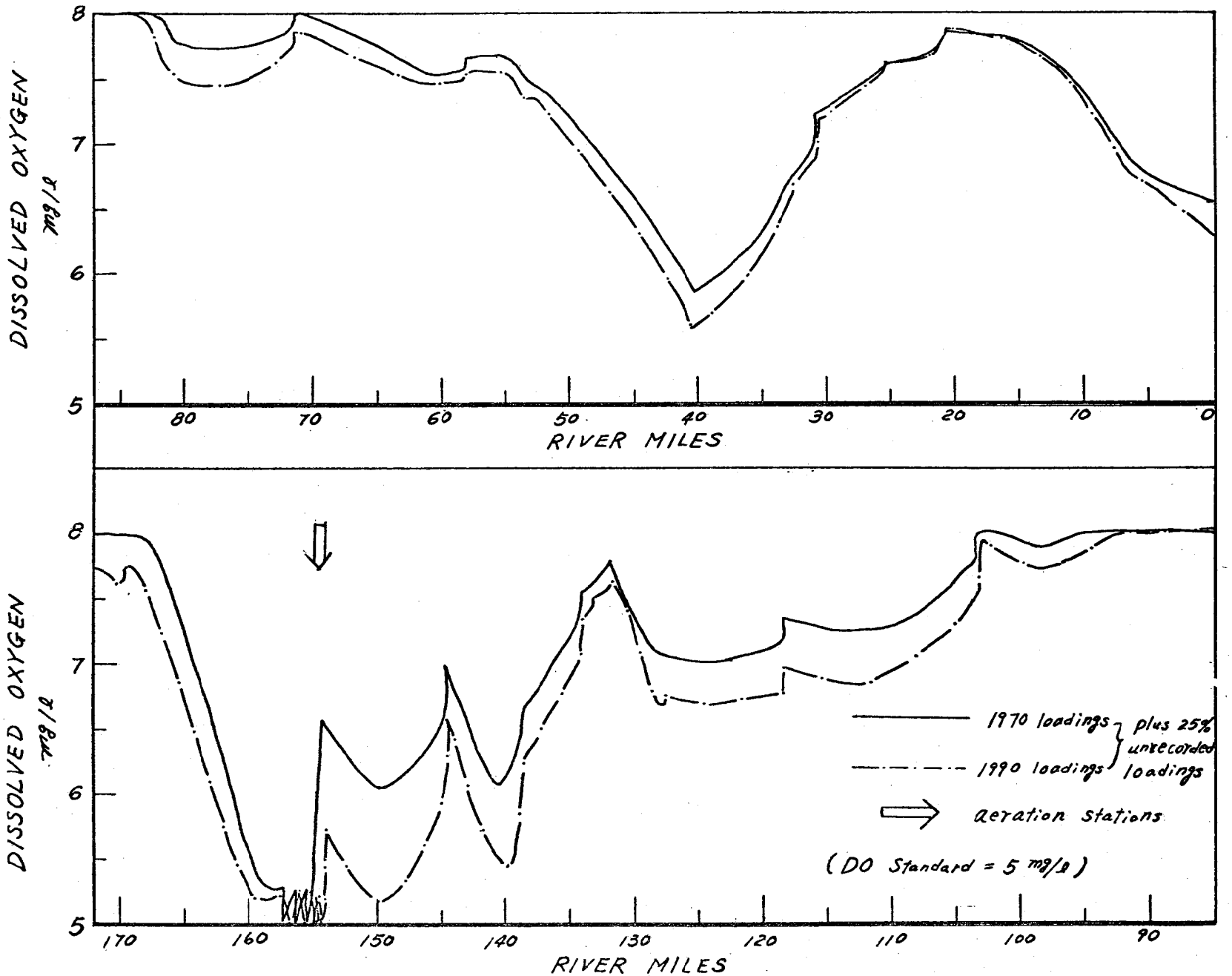


FIG. 5 DO PROFILES WITH SECONDARY TREATMENT PLUS AERATION

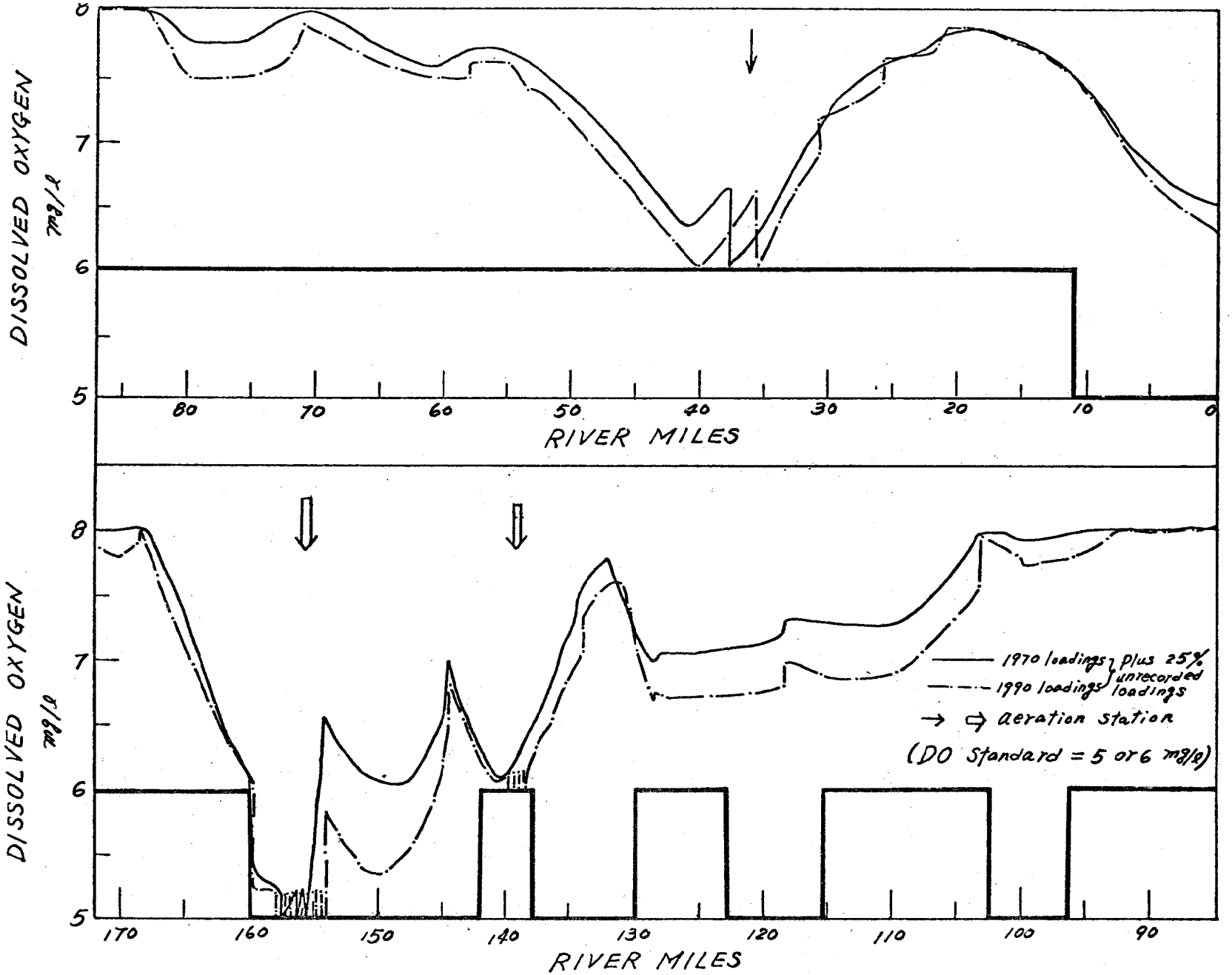


FIG. 6 DO PROFILES WITH SECONDARY TREATMENT PLUS AERATION

required at each aerator. In most cases excavation in the river bed would be necessary to provide a satisfactory operating depth. Electric service costs have been estimated assuming a distance of 500 feet from the power source. Capital recovery costs were estimated on the basis of a six per cent interest rate and a life of 15 years. It is assumed that each aerator would be a separate installation, located a significant distance from the next, but close enough that maintenance personnel could drive readily from one to another. Allowance has been made for aerators to be removed from the river each fall for overhaul and protection from ice. Electric energy estimates were based upon continuous operation during three months of each year and half-time operation for five additional months. Personnel costs included provision for a daily visit to each site, with one man to inspect, operate and maintain a number of installations, and with an additional allowance of 3% annually of total material costs for more important repairs. There has been no allowance for overhead charges or supervision by any organization above the field level.

Estimated costs are given in Table 2 for systems of 5, 20, and 35 aerators.

Thus, on the assumption of 1970 waste loads and no unrecorded waste loading, and the further assumption of secondary treatment sufficient to remove 80% of carbonaceous oxygen demand and 30% of nitrogenous oxygen demand, a system of five aerators would be required in addition to the treatment, in order to meet desired stream flow standard of 5-6 mg/ℓ of dissolved oxygen. The capital costs for this aeration system would be about $231,000, and the operation and maintenance costs $53,000 annually.

TABLE 2

ESTIMATES OF COST OF AERATOR SYSTEMS

No. of 75 hp Aerators	5	20	35
Mechanical aerator electrical equipment installation, excavation, etc.	$193,000	$776,000	$1,359,000
Contingencies and engineering	$38,000	$155,000	$272,000
Total capital costs	$231,000	$931,000	$1,631,000

Annual operation and maintenance costs	$53,000	$167,000	$294,000
Capital recovery costs 6% - 15 years	$23,000	$96,000	$168,000
Total annual economic costs	$76,000	$263,000	$462,000

9. Discussion

Estimates could readily be made of costs of aeration systems which would achieve the same dissolved oxygen levels without secondary waste treatment. However, it would not be wise to consider such an aeration system in the present state of knowledge. Secondary waste treatment is necessary not only in order to reduce oxygen demand, but also in order to remove suspended solids, and various other pollutants, some of which may be unknown. Therefore, an instream aeration system should be considered as supplementary to secondary waste treatment, but not as an alternate to it. In the Merrimack River, the main need appears to be for secondary waste treatment.

However, on the Merrimack River the assumption which has been made that there are no unrecorded waste loads is manifestly incorrect. A mass balance analysis would undoubtedly show that a considerable proportion of mean waste loadings consists of unrecorded BOD [from both point and non-point sources], and that most of this unrecorded BOD will not, as assumed, be subject to 80% reduction in the event of a general program of secondary waste treatment (6). Such unrecorded loadings are particularly marked after rainfall, during periods of flow greater than those for which the water quality model was derived. Therefore, alternative estimates allowing for inclusion of 25% of unrecorded BOD have been prepared and are given in Table 1. However, without a basin-wide mass balance analysis, the amount of unrecorded BOD to be allowed for is purely conjectural.

In order to determine the economic feasibility of instream aeration systems such as those referred to above, cost estimates should be made of incremental waste treatment programs which would be directly comparable in their effects. However, in order to be

realistic, the amount of unrecorded waste loading in the basin should be estimated, and should be excluded from the waste loading assumed to be reduced by treatment.

REFERENCES

(1) Camp, T.R., "Water and Its Impurities," Reinhold Publishing Co., New York, 1963.

(2) FWPCA, Report on Pollution of the Merrimack River - and Certain Tributaries, August, 1966.

(3) O'Connor, D.J., Stream and Esturine Analysis, Manhattan College, New York, Summer Institute notes.

(4) O'Connor, D.J., J.P. St. John and D.M. DiToro, "Water Quality Analysis of the Delaware River Estuary," Journal SED, ASCE, 94, SA6, December 1968, pp. 1225-1252.

(5) Chapter 3 in Whipple, et. al., "Instream Aeration of Polluted Rivers," Water Resources Research Institute, Rutgers University, 1969. [Out of print, obtain from Clearinghouse, No. PB 192637.]

(6) Whipple, W. Jr., "BOD Mass Balance and Water Quality Standards," Water Resources Research, June 1970, pp. 827-837.

(7) Ludzack, F.J. and N.B. Ettinger, "Controlling Operation to Minimize Activated Sludge Effluent Nitrogen," Journal, WPCF, Vol. 34, No. 9, 1962.

☆ U.S. GOVERNMENT PRINTING OFFICE: 1971—700—195/6