EPA-670/2-74-080
DECEMBER 1974

Environmental Protection Technology Series

STUDY OF AN INTEGRATED POWER, WATER AND WASTEWATER UTILITY COMPLEX

National Environmental Research Center
Office of Research and Development
U.S. Environmental Protection Agency
Cincinnati, Ohio 45268

EPA-670/2-74-080
December 1974

STUDY OF AN INTEGRATED POWER, WATER AND WASTEWATER UTILITY COMPLEX

By

New York State Atomic and
Space Development Authority
New York, New York 10017

Project No. 17080 HHV
Program Element 1BB043

Project Officers

John R. Trax and Patrick Tobin
U.S. Environmental Protection Agency
Washington, D. C. 20460

NATIONAL ENVIRONMENTAL RESEARCH CENTER
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

REVIEW NOTICE

The National Environmental Research Center -- Cincinnati has reviewed this report and approved its publication. Approval does not signify that the contents necessarily reflect the views and policies of the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

Man and his environment must be protected from the adverse effects of pesticides, radiation, noise and other forms of pollution, and the unwise management of solid waste. Efforts to protect the environment require a focus that recognizes the interplay between the components of our physical environment--air, water, and land. The National Environmental Research Centers provide this multidisciplinary focus through programs engaged in

- studies on the effects of environmental contaminants on man and the biosphere, and
- a search for ways to prevent contamination and to recycle valuable resources.

Distillation has long been known as a method for producing very pure water. This report discusses how distillation of wastewater with heat from an electric power plant might be utilized for producing reuseable water.

A. W. Breidenbach, Ph.D.
Director
National Environmental
Research Center, Cincinnati

ABSTRACT

This study evaluates, technically and economically, a new approach to siting power generation, wastewater treatment and water supply facilities. It is concluded that the integrated facility results in more efficient utilization of land and water resources, produces a net reduction in undesirable process effluents, and achieves at a reduced cost many of the environmental quality goals sought today. In particular, the use of waste heat for the beneficiation of wastewater treatment was determined to be sufficiently promising to merit further investigatory research.

The integrated facility studied will supply 1000 Mw of electric power at 9.1 mills/kw-hr, will provide secondary treatment for 50 MGD of wastewater for 15¢/1000 gal., and will produce 47.5 MGD of high quality potable water for approximately 62¢/1000 gal. utilizing low quality steam and waste heat.

A three phase follow-on research and demonstration program is defined and is directed toward the development of the further design and performance information necessary to permit the undertaking of full scale integrated facilities.

This report was submitted in fulfillment of Project Number 17080 HHV by the New York State Atomic and Space Development Authority under the sponsorship of the Environmental Protection Agency.

TABLE OF CONTENTS

FIGURES

NO

TABLES

TABLES

SECTION I

CONCLUSIONS

Conceptual designs were developed to determine the technical and economic feasibility of integrating a 1000 Mwe nuclear power plant, a 50 MGD wastewater plant utilizing waste heat to enhance the treatment process, and a 47.5 MGD distillation plant producing potable water from wastewater, based on a reference site located on the north shore of Long Island near Riverhead in Suffolk County, New York.

Detailed evaluations of installing and operating the multiple utility complex at that location, and of utilizing waste heat to elevate the temperature for processing wastewater resulted in the principal conclusions that:

(a) The integration of facilities for nuclear power production, wastewater treatment and distillation of secondary treated wastewater is technically feasible and economically attractive.

(b) The use of power plant waste heat to improve the efficiency of the unit operations in primary and secondary treatment of wastewater is indicated to be sufficiently practicable to warrant further investigation and demonstration.

It was determined that the reference design integrated facility would cost $386 million and would provide electrical power for 9.1 mills/Kw-hr, wastewater treatment for 15.1¢/1000 gal. (with no effluent discharge) and potable water for 62.4¢/1000 gal. Separate facilities having the same capacities would cost $284 million, $25.5 million, and $83.8 million for the power, wastewater treatment, and desalination facilities, respectively.

The savings of $7,300,000 in capital cost and $179,000 in annual operating cost (which does not give effect to the additional fuel cost savings available) are offered by the integrated facility as compared to the individual facilities it replaces. In addition the integrated facility accomplishes environmental benefit which the conventional, individual facility alternative does not -- its wastewater treatment operations do not release liquid effluents to the environment.

The cost of distilling secondary treated wastewater should be less than the cost of desalting seawater in a combination electric power/seawater distillation plant in view of the lower concentrations of dissolved material in the treated wastewater. Moreover, the distillation process is an effective method of ammonia control, and the use of this process should preclude the build up of nitrates in recycle or recharge systems. Because of the addition of heat, ammonia can be removed on a year-round basis and is not subject to the limitations of conventional ammonia strippers. It should be noted that the distillation of raw and primary treated wastes is not considered to be technically or economically feasible based on the present information.

By using waste heat to increase wastewater temperature from $65^{\circ}F$ to $93^{\circ}F$, the performance of the grit chamber, the primary and secondary clarifiers, the aeration tanks and the sludge thickeners would increase in efficiency by 18 to 30 percent. The improved performance of these processes would result either from improved settling rates due to the decrease in density and viscosity of the wastewater, or from increase in the biological activity in the organic waste assimilation processes, or from a combination thereof.

Heating of the untreated wastewater to $93^{\circ}F$ by use of steam from the low pressure stages of the distillation plant, injected by means of a barometric leg condenser -- the conservative approach selected for this study -- would result in some additional capital and operating costs since the distillation and wastewater plants must accommodate the 1.4 MGD of condensate added to the wastewater. The estimated added cost of $1,990,000 for this arrangement is of the same magnitude as the $1.5 million saving which could be realized through the reduction in size of a heated wastewater plant.

The potential economic advantage for heating wastewater lies in demonstrating (either by verifying the capability of existing equipment, or by advancing the technology) that surface type heat exchangers are capable of realizing heat transfer coefficients in the range of 150 to 350 Btu/hr-ft^2-$^{\circ}F$, and that such exchangers can be installed in a cost range of $12 to $27 per square foot. Under these conditions, surface exchangers utilizing wastewater could be used for cooling the product water, thereby eliminating the 1.4 MGD of additional processing capacity in the wastewater treatment and distillation plants and accordingly reducing the cost of these facilities by the $1,990,000 indicated above. In addition, the product water/seawater exchanger which is estimated to cost $1,869,000 would be eliminated. The magnitude of the saving realized by substituting product water/wastewater exchangers will depend on the heat transfer coefficients and heat exchanger cost reductions achieved in the demonstration project.

SECTION II

RECOMMENDATIONS

The economic and technical results of this study indicate that a research and demonstration program should be undertaken toward the development of the technology, design and performance information that will be required to effect future full scale integration of the facilities for power generation, wastewater treatment, and water supply. A program addressed to these objectives would be conducted in three phases:

Phase I	Laboratory Screening Studies and Component Tests
Phase II	System Performance Tests and Demonstrations
Phase III	Integrated Pilot Scale Demonstration Tests

The work to be performed under Phase I would consist of two separate but related studies -- a wastewater laboratory screening study and a heat exchanger component testing study.

The laboratory screening studies to be conducted in Phase I would be designed to,

(a) Determine actual performance characteristics of activated sludge treatment operations at the elevated temperatures considered in this analysis.

(b) Evaluate the qualitative changes in the nature of the activated sludge treatment process at elevated temperatures.

(c) Determine the performance characteristics of non-biological treatment operations at elevated temperatures, for example, chemical treatment and chemical sludge disposal.

Laboratory screening results would be analyzed in terms of performance and economic implications to arrive at a priority ordering of unit operations for pilot plant verification.

The heat exchanger component testing program would be designed to contribute to the development of surface condensers and heat exchangers capable of transferring low grade energy to wastewater from power plant exhaust steam condenser cooling water or distillation plant product water reliably and economically. To accomplish this objective, tests would be designed to establish,

(a) fouling factors and overall heat transfer coefficients;

(b) the corrosion resistance of candidate wastewater heat exchanger materials;

(c) methods and procedures for the reduction and removal of scale.

The system performance tests and demonstrations of Phase II planned on the basis of Phase I laboratory screening results and component tests, would be undertaken to verify the performance characteristics at elevated temperatures of selected wastewater treatment processes. The system performance tests would provide an opportunity for verification of the suitability of heat transfer equipment identified in the previous phase. Performance data would be developed on the ability of distillation plants to remove ammonia and produce water of high quality reliably and continuously from treated wastewater. These tests would also establish requirements for post-distillation treatment and plant product water quality control. Finally, the tests and demonstrations would be analyzed to identify performance and economic aspects relevant to the development of specific process flow sheets and selection of heat transfer and other equipment for the Pilot Scale Demonstration.

In Phase III heated wastewater treatment followed by distillation would be demonstrated at the site of an existing power plant and wastewater treatment plant having a capacity of up to 4 MGD.

SECTION III

INTRODUCTION

In many areas of the United States, communities are experiencing difficulty in providing adequate supplies of water to meet the needs of a rapidly growing population. This growth and an increased per capita water usage are reflected in an increasing need for sewage treatment facilities as well as for storm water and domestic sewage conveyance systems all of which accelerate the depletion of potable water supplies in regions which are dependent upon ground water.

In the urban areas of this country, the rate of water extraction from our rivers, lakes, and aquifers has exceeded the rate of replenishment through the natural hydrologic cycle. Suburban areas in Long Island are experiencing a growth rate substantially above the national average at a time when groundwater resources, the only source of supply, is deteriorating in quality and diminishing in quantity. Moreover, the rapid rate of growth has created a concomitant demand for additional power generation.

Recently, considerable public attention has been focused on the environmental impact associated with the siting of major facilities such as power plants, reservoirs, and wastewater treatment plants. Not only is the selection of sites becoming more expensive due to the environmental studies required and the inclusion of more stringent environmental safeguards, but in many areas of the country, the prospects for finding facility locations are extremely limited. Suburban development, industrial activities, recreation facilities and other competing uses for land have already claimed many prime facility sites.

Historically, organizations responsible for siting major public and private facilities have proceeded in an independent manner. For example, sanitary districts seek and acquire sites for wastewater treatment plants, water utilities locate reservoirs and treatment plants, and electric utilities acquire sites for future generating stations. In each case, the specific environmental impact of waste products must be considered. Federal, state and local standards now regulate the quality of air and water discharges from the separate facilities. The cost of adequate effluent treatment is passed on to the eventual consumer of the product or services.

The hypothesis that a new approach to utility siting, that of jointly siting power generating, wastewater treatment and water supply facilities, results in a more efficient utilization of land and water resources, a net reduction in undesirable process effluents, and the accomplishment of many of the new environmental quality goals at a reduced cost when compared to separate facilities, is the subject of this study.

Thus, the overall objective of the project was to determine the feasibility of combining electric power production, wastewater treatment, and potable water supply utilities, and to evaluate the economic and environmental advantage of such a combined facility for treating wastewater for reuse as an alternate water supply. The specific objectives were:

1. To define the technical considerations involved in the integrated siting of power generation, wastewater treatment, water supply and distillation facilities.

2. To select an illustrative site and prepare a design of an integrated facility.

3. To determine the effectiveness of the integrated facility in meeting future water demand.

4. To establish the principal design and performance aspects associated with integrated utility complexes.

5. To establish the economic benefits and costs of a full scale integrated facility.

6. To identify the beneficial effects of heat addition in wastewater processing.

7. To define the necessary development and demonstration projects.

In order to determine the benefits associated with an integrated facility, an engineering and economic analysis was performed. The power production, wastewater treatment and water supply systems were analyzed to identify the most effective mode of overall integration. The integrated facility was compared to the conventional method of using individual facilities for power, wastewater and water services. Since the design and economic evaluation of both the integrated facility and the individual facilities was strongly site dependent, a specific location was identified.

The New York State Atomic and Space Development Authority undertook this study pursuant to its responsibilities for conduct and fostering the use of atomic energy for productive purposes. The Authority was joined in this work by environmental science and engineering firm of Quirk, Lawler and Matusky, which was responsible for the analytical and design work and economic analysis of the wastewater treatment systems of the integrated plant discussed in Sections V, VIII and X, as well as for the determination of the effect of elevated temperatures upon wastewater treatment processes work presented in Appendix A and by Hittman Associates, Inc. The latter firm reviewed and assessed the information presented in Section IV, performed the analytical design and economic calculations for the nuclear power facilities presented in Sections V through VII, the distillation plant design presented in IX, those portions of the overall plant integration and evaluation pertinent to their work, as presented in Section X, and the evaluation of engineering and economic factors affecting addition to wastewater treatment processes, as presented in Appendix B.

Section XI, which sets forth the development and recommended further demonstration projects, was prepared jointly by Quirk, Lawler and Matusky, Engineers and Hittman Associates.

SECTION IV

THE REGION AND ITS RESOURCES

REGIONAL GEOGRAPHY

Suffolk County, the easternmost county on Long Island, covers a land area of approximately 920 square miles. The county is bounded on the north by Long Island Sound, on the east and south by the North Atlantic Ocean, and on the west by Nassau County. Its Nassau County boundary is within 15 miles of the city limits of New York City and within 30 miles of central Manhattan.

The county is approximately 86 miles long and 21 miles wide at its widest point which is along its western boundary. The major land mass extends east-northeast from the Nassau County line for 42 miles to Riverhead; east of Riverhead the land mass is bifurcated into peninsulae extending eastward and separated by a series of bays, all shown on Figure 1, which also identifies the reference site location. The north fork, extends approximately 28 miles east of Riverhead; the larger southern fork is approximately 44 miles long and terminates at Montauk Point, the easternmost point of New York State.

POPULATION PROJECTIONS

Suffolk County is composed of 10 towns ranging in size from 11 square miles (Shelter Island) to 252 square miles (Brookhaven). Until recent years, the county was almost entirely agricultural in nature, but now agricultural uses predominate only in the eastern areas. In recent years the increasing population encroachment from Nassau County and extension of the continuing growth surrounding the New York metropolitan area have resulted in rapid increases in population and population density in the western part of Suffolk County. A 70 percent increase in the total population of the county, from 667,000 to 1,127,000, was observed between 1960 and 1970 (1). Approximately 97 percent of this increase was observed in the five western towns which, in 1970, were found

Figure 1. Reference Site Location

to have a total population of approximately 1,043,000, or approximately 93 percent of the total in the county situated on approximately 37 percent of the total county land area (2).

Although the population growth rate in the United States has shown a decrease over the last decade and this trend is expected to continue, the population of Suffolk County is forecasted to grow at rates substantially higher than the national average. The dominant factor in this growth will be the increasing scarcity of suitable land for new home sites closer to the New York metropolitan area, with the most rapid growth occurring in the westernmost towns.

TABLE 1

LONG ISLAND POPULATION PROJECTIONS
(In Thousands)

Year	Nassau County (Ref.4)	Suffolk County (Ref.3)	Suffolk County (Ref.4)
1970	1429	1127	1127
1975	1565	1276	1328
1980	1651	1515	1570
1985	1723	1753	1825
1990	1743	1978	2190
1995	1791	2198	2360
2000	1836	2379	2500
2005	1879	----	2645
2010	1928	----	2790
2015	1977	----	2920
2020	2021	----	3050

Table 1 presents population projections for both Suffolk County and adjacent Nassau County. These data are based primarily on a report commissioned by the Nassau-Suffolk County Regional Planning Board (4). New York State data (3) for the years 1970-2000 for Suffolk County are also included for comparison purposes. Close agreement may be noted, with slightly lower population growth estimated by the state. Comparable data were also obtained from a report prepared in evaluation of sewerage needs for eastern Suffolk County (2). This report indicated a permanent population of approximately 2,046,000 in Suffolk County in the year 1990 and 2,960,000 in the year 2020. As there is no significant conflict between these projections, the data from the County Regional Planning Board were utilized throughout this report in the development of further analyses of power and water requirements based on population growth.

In addition to the overall county population information, estimates of the population in each of the ten individual Suffolk County towns for the years 1970 through 2020 were extracted from Reference 4 and are presented in Table 2. Current New York State projections for the years 1970 to 2020 are presented parenthetically for comparison purposes (3). Each town covers an extensive area and is comprised of a number of small villages and hamlets.

WATER RESOURCES, SUPPLY AND DEMAND

Groundwater constitutes virtually all the water supply available to Suffolk County. Of the 44 inch average annual precipitation, approximately two percent is lost through direct runoff and 48 percent returns to the atmosphere via evapotranspiration, leaving approximately 50 percent of the average annual rainfall available for groundwater recharge. This natural recharge corresponds to about one million gallons per day per square mile.

The surface water supplies are quite meager since there are no major lakes within its boundaries. The largest body of water is Lake Ronkonkoma which covers an area of only 245 acres, and most of the streams are estuarine; that is, salty in their lower reaches where they flow into bays or directly into the ocean or sound. It is estimated that 95 percent of the fresh water streamflow is due to discharge from the groundwater reservoir, with the rest being direct runoff (5).

TABLE 2

SUFFOLK COUNTY TOWN POPULATION PROJECTIONS
(In Thousands)

Town*		1970		1980		1990		2000		2010	2020
Babylon	(10/5)	200	(204)	262	(243)	283	(281)	292	(327)	298	300
Huntington	(18/1)	200	(201)	280	(252)	315	(311)	325	(354)	330	333
Islip	(16/3)	276	(279)	358	(353)	378	(400)	385	(456)	400	405
Smithtown	(9/0)	115	(114)	160	(167)	170	(208)	175	(239)	179	181
Brookhaven	(25/1)	235	(244)	375	(395)	700	(626)	840	(800)	890	920
Southampton	(12/0)	40	(36)	65	(44)	116	(59)	170	(76)	224	267
Riverhead	(1/0)	19	(19)	30	(26)	63	(39)	130	(53)	198	260
East Hampton	(3/0)	17	(11)	39	(14)	70	(20)	110	(26)	142	165
Southold	(3/0)	18	(17)	27	(20)	42	(30)	73	(41)	120	214
Shelter Island	(1/0)	2	(2)	3	(2)	3	(4)	4	(7)	8	12
Total Suffolk County		1127	(1127)	1600	(1515)	2190	(1978)	2500	(2379)	2800	3050

*-The first number in parenthesis after the town indicates the number of villages or hamlets within the town and the second number indicates the number having a population of over 15,000.

The earliest public water supply systems on Long Island depended on surface water for their source of supply. When the original Brooklyn system was completed in 1862, the supply consisted entirely of surface water with gravity flow distribution. In 1872 the first Brooklyn pump stations were installed. This second water supply system on Long Island augmented the original system by pumping water from ponds.

As the meager surface water supplies diminished, the growing population turned to groundwater for its increased needs. In 1874 the first groundwater for public supply was used on Long Island by Long Island City. By 1880 Brooklyn also began exploiting groundwater. Thus, by 1902 there were 120 MGD being used for public water supply on Long Island, 65 MGD from surface sources and 55 MGD from the ground. Brooklyn alone required 85 MGD, and obtained 60 MGD from surface supplies and 25 MGD from wells.

As the western part of Long Island developed rapidly, the increased impermeability of the urbanized areas reduced the infiltration capacity, while establishment of sanitary and storm sewers resulted in removal of water which would normally recharge the aquifers. As a result, groundwater mining became commonplace. In the 1930's, excessive pumpage in Brooklyn lowered local groundwater levels to as much as 35 feet below sea level. This condition caused salt water to contaminate a large portion of the groundwater reservoir, and forced a suspension of further withdrawals. Recovery from this condition of depletion is extremely slow due to the high degree of urbanization and the construction of sewer outfalls to the ocean.

With the eastward migration of population on Long Island, the attendant water supply problems also move eastward. Recent estimates (4) of consumptive use and of the groundwater resources of Nassau County show that there is an average annual decrease in the groundwater storage in the county of 5 MGD due to groundwater mining. Consumptive use is defined as the rendering of water unavailable for reuse until such water has passed through the precipitation part of the hydrologic cycle. Consumptive use includes water discharged through sewer systems to the sea, infiltration of groundwater into the sewer system, evapotranspiration following irrigation and lawn sprinkling, and the relatively small amount consumed and evaporated during general use. Figure 2 shows the water budget for Nassau County (5).

A comprehensive study of the public water supply and requirements of Suffolk and Nassau Counties was recently completed and published (4) which indicated a present per capita water usage in Suffolk County of approximately 100 gallons per day (gpd), and a future per capita usage of approximately 150 gpd by the year 2020. On the basis of the present and projected population, the requirement for water exclusive of agricultural uses is estimated to be approximately 120 MGD at the present time and approximately 450 MGD by the year 2020

In general, Suffolk County has not felt the acute water supply problems caused by development and increased population. Its water supply is sensitive to precipitation levels and during periods of extended drought of the type which occurred in the mid-1960's. Such problems could be quite serious for many of its communities. Figure 3 shows the present generalized water budget for Suffolk County. As the population of the county grows, however, large increases in the amount of human consumptive use will occur, due, in part, to the general increase in population, in part to the increase in the amount of sewered area, and in part to the general trend toward increased per capita use which accompanies increased personal income. Even if no additional sewers are constructed, consumptive use is expected to approach 100 MGD by the year 2020 (3, 4).

Although the Suffolk County water budget for the year 2020 is somewhat a matter of conjecture at this time, some generalized conclusions can be drawn with the aid of Figure 4. Population estimates place the population of Suffolk County in the year 2020 at about 2 1/2 times the present population,

Assuming a conservative increase in the amount of storm sewer systems, direct runoff would probably increase to at least 100 MGD. With evapotranspiration remaining constant, the amount of precipitation available for groundwater recharge would, necessarily, decrease. Human consumptive use is shown to increase significantly by the year 2020, as a result of a more than fourfold increase in the area serviced by sanitary sewers.

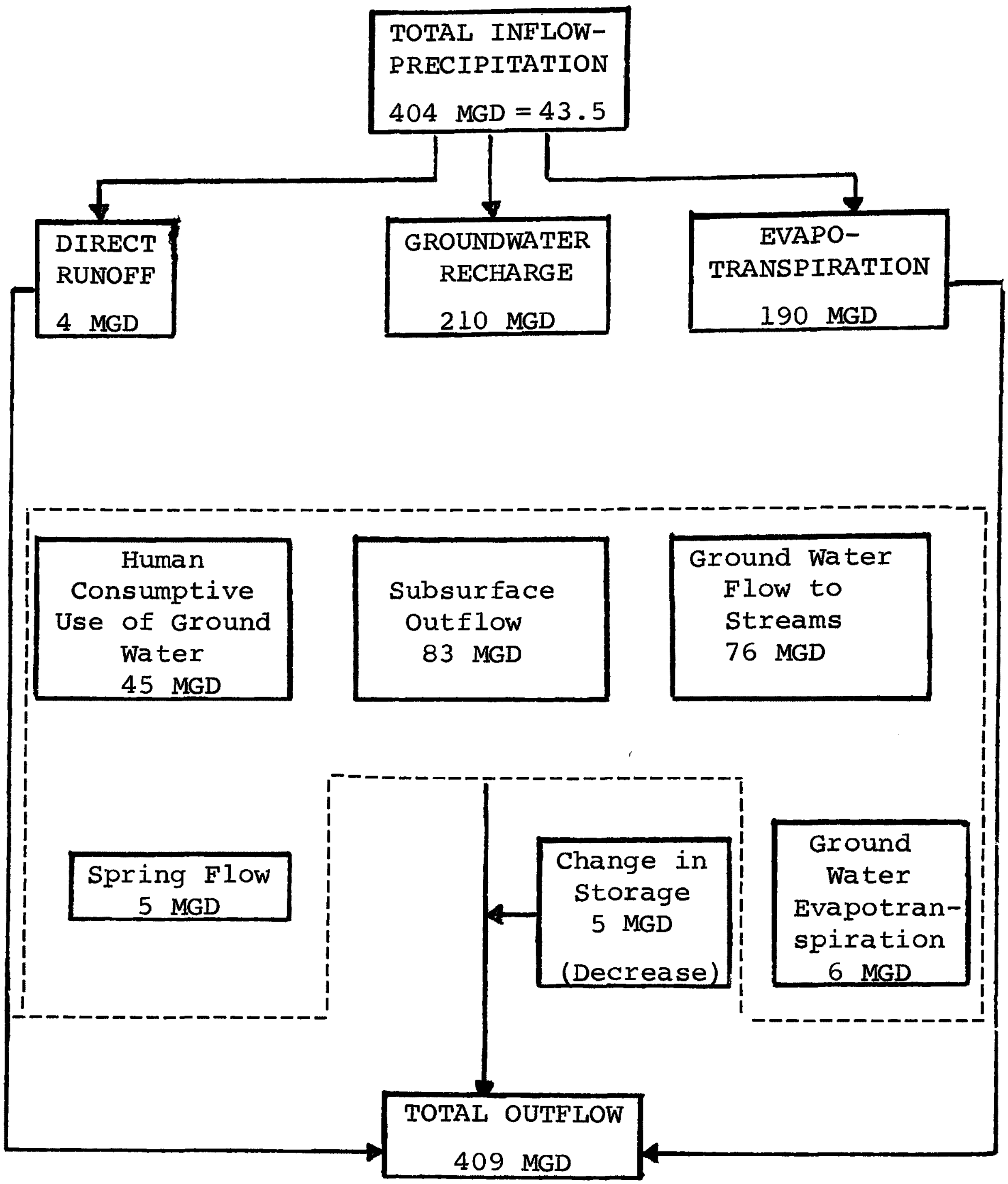


Figure 2. Nassau County Water Budget (Based on Ref. 5)

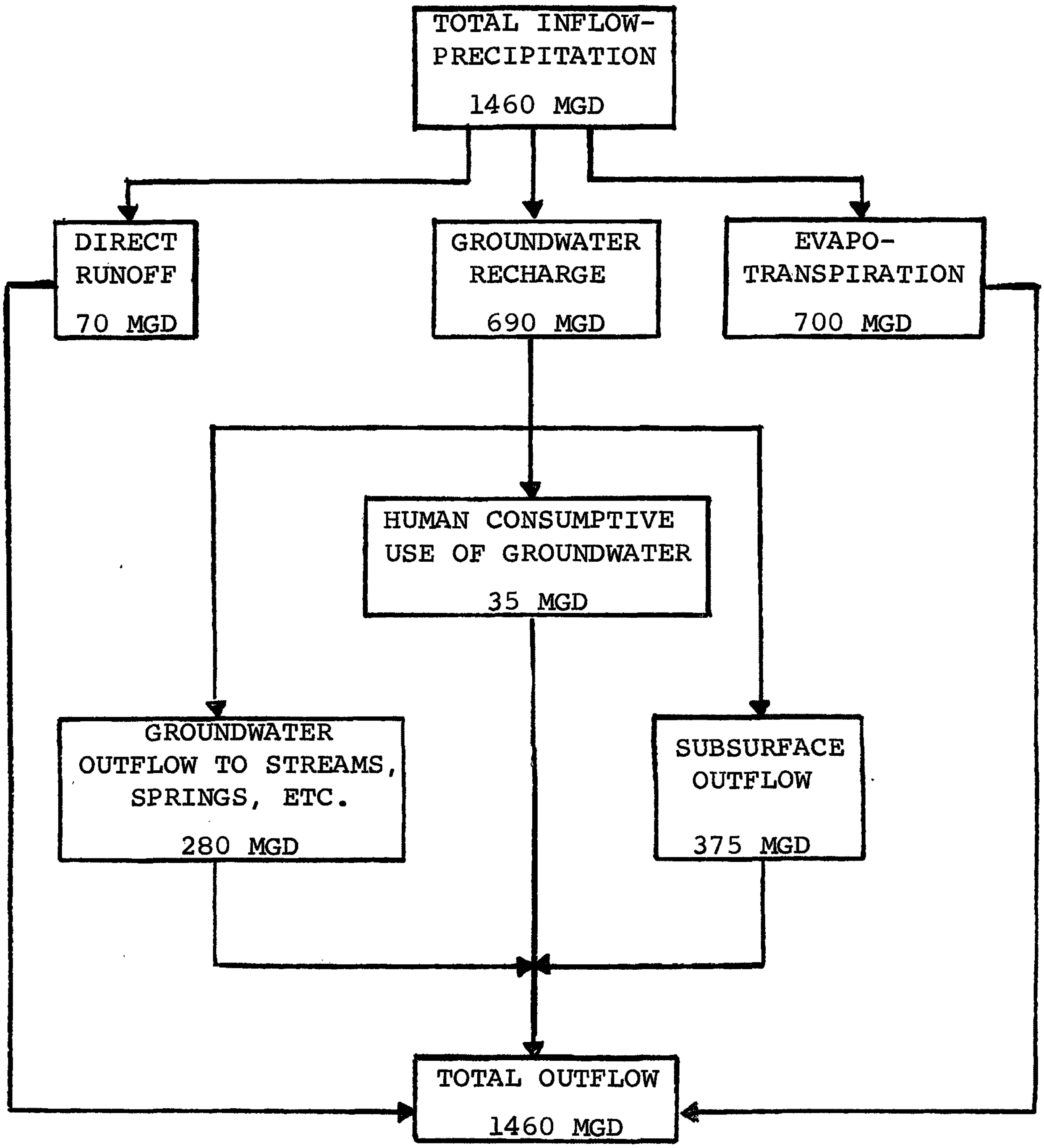


Figure 3. Present Suffolk County Generalized Water Budget

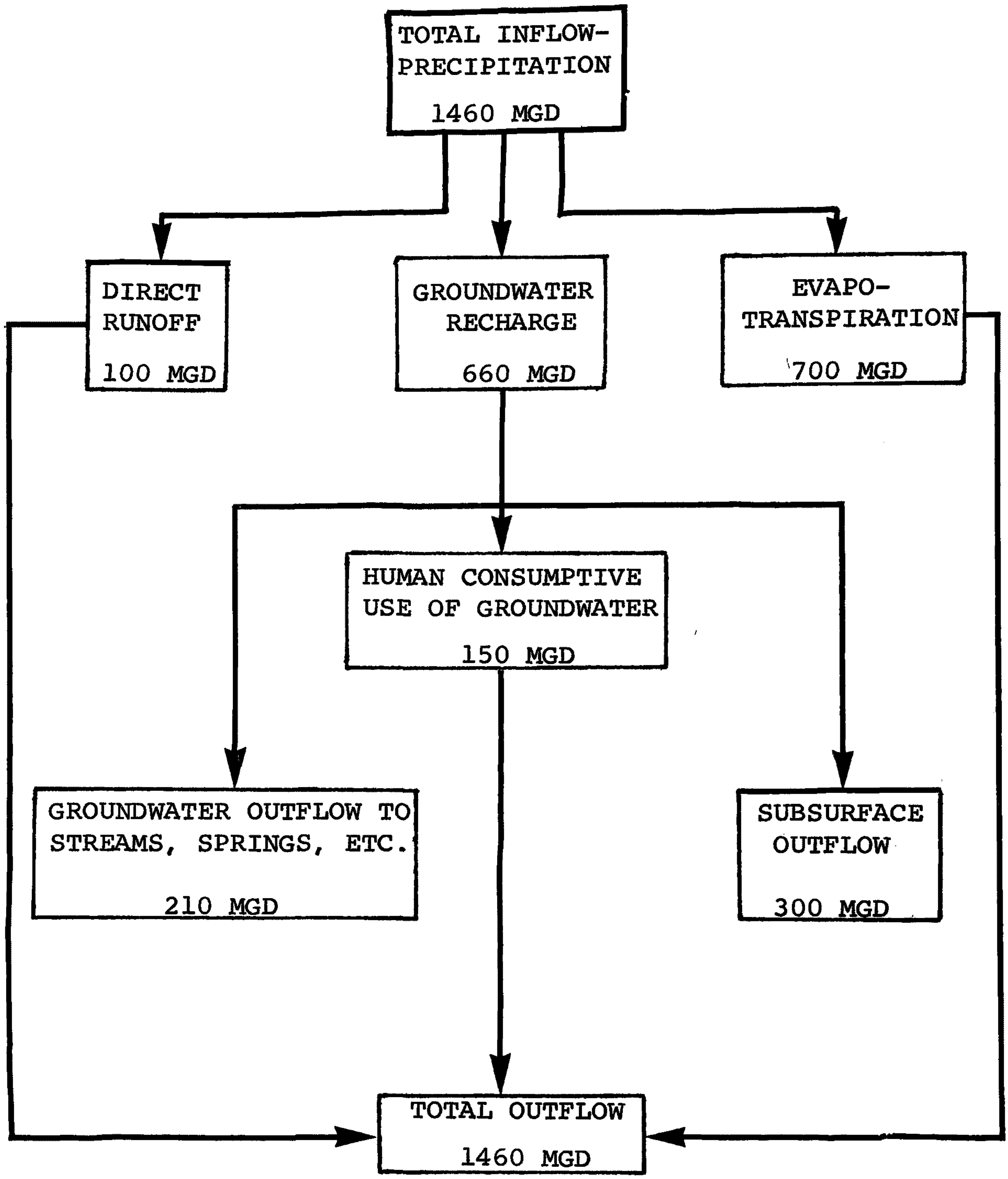


Figure 4. Suffolk County Generalized Water Budget for the Year 2020.

A comparison of Figures 3 and 4 shows that by 2020 the projected increase in consumptive use of water will reduce subsurface outflow and groundwater outflow to streams and springs. This increased consumptive use could also result in depletion of the groundwater reserve as is occurring in Nassau County. Thus, the 20 percent decrease in subsurface outflow and the 25 percent decrease in the groundwater outflow to streams and springs would have a profound effect on the groundwater balance. Accordingly, streamflow will show a marked decrease, especially during periods of drought, as was evidenced in Nassau County during the drought period of mid-1960's by the zero flow condition recorded in some streams in areas that were predominantly sewered (4). As subsurface outflow and streamflow decrease or as groundwater reserves are directly depleted, a marked increase in the amount of saltwater encroachment will also occur, further deteriorating the water supply situation.

The safe water yield in Suffolk County in 1957 was estimated to be approximately 501 MGD (7), which would, if such yield could be sustained in the face of the currently projected development, meet the average per capita demand of 150 gpd for all purposes for a population of 3,340,000. Considering the uncertainty in such yield and population projections, the 13-17 percent increase in summer population over that of winter, and the recognized variation in peak daily and seasonal demands from the annual average, the sustained safe groundwater yield may well be exceeded by the demand during the period under consideration (1970-2020). Even if the county-wide yield is not exceeded by the demand, the anticipated large imbalance in the population distribution will necessitate an extensive water distribution system providing either groundwater from the central portion of the county or water from sources outside the county for the western towns. It must be noted, however, that Suffolk County is the only county on Long Island where demand does not already exceed supply. Consequently, if an additional source of water for the county is to be found it must be imported from the mainland (i.e., from upper New York State through the New York City metropolitan water system), through desalting of seawater, or through recovery and reuse of wastewater which would otherwise be discharged to the Atlantic Ocean or to Long Island Sound.

An additional consideration in the water supply picture of Suffolk County is the continued pumping and groundwater mining in adjacent Nassau County. If the water levels in Nassau continue to decline, Suffolk County will eventually feel the impact of reduced groundwater storage since hydrologic conditions do not respect political boundaries.

If water now lost through human consumptive use can be recycled through wastewater treatment and distillation, a significant step will have been taken toward meeting future water supply needs. Probably the most critical area within Suffolk County is the north fork, containing the town of Southold and about half the town of Riverhead. Here, the fresh groundwater reservoir is restricted to relatively thin lenses in the glacial or Upper Pleistocene deposits. Withdrawal of water from these lenses is thus limited to shallow wells of low capacity and, therefore, well spacing and distance from the sea are of great importance.

The towns of Riverhead and Southold, closest to the proposed site, have present water requirements of approximately 3 MGD each. These requirements are anticipated to increase by a factor of three to four by the year 2000 and to further increase by an additional factor of approximately three in the following two decades. Thus, depending on the population growth, the water requirements in each of these towns by the year 2000 and 2020 are expected to be on the order of 30 MGD and may exceed 50 MGD during peak periods.

The projected population increases and groundwater withdrawals, the greater use of sewers, the additional runoff and higher evapotranspiration due to development may be expected to lower the area's groundwater level and quality. Seasonal demands currently result in heavy withdrawals of groundwater, such as in summer months, when both agricultural and summer resort residential demands increase significantly. Agricultural pumpage has already created local problems of saltwater intrusion.

Moreover, during periods of drought such as that experienced in the 1960's, local lowering of the water tables occurs. The low groundwater levels and reduced surface flow under such conditions have resulted in some water quality problems locally. In addressing this projected need, the water produced from the proposed integrated facility can be considered as a new source of fresh water. The wastewater from which it is derived would normally be discharged to the sea. By recycling this water, the supply of fresh water in the region is augmented and the other sources of fresh water are conserved and protected. The product water from the plant could be used to recharge the groundwater aquifers or for industrial purposes. Alternatively, distilled water product could be made available for distribution. In this case essentially 100 percent of the recovered water would be put to use, whereas if used for groundwater recharge approximately 50 percent would be lost. In any case, the quality of the water produced by distillation would be suitable for either recharge or direct reuse.

TABLE 3

PROJECTED WATER USAGE
FOR THE YEARS 1990 AND 2020
FOR DISPOSAL DISTRICTS 11 AND 13*

District	1990 Population	1990 Total Water Usage (MGD)	2020 Population	2020 Total Water Usage (MGD)
11	229,000	59.6	357,000	61.2
13	84,000	19.0	218,000	29.5
Total	313,000	78.6	575,000	90.7

*-Based on Reference 4.

Table 3 shows the projected total water usage for Disposal Districts 11 and 13, which would supply wastewater to the integrated facility for the years 1990 and 2020. If an integrated facility should be operational by 1990, about 61 percent of the average annual water requirements for these districts could be recovered by the integrated plant. By 2020, when the average annual requirement has increased to 90.7 MGD, the plant recovery would represent 53 percent.

A portion of the product water would be used in Suffolk County to reduce withdrawals in critical areas and to provide adequate reserves for drought periods through reduction in demand on groundwater.

In the near-term one potential use for the product water would be in neighboring Nassau County, where there exists a groundwater mining situation such that new sources of fresh water in the amount of 94 MGD in 1990 (6) will be required to prevent saltwater infiltration and/or encroachment. Exporting water to Nassau County will also benefit Suffolk County since the lowering of the groundwater levels in Nassau County would cause a lowering of the groundwater in Suffolk County. Secondary treated wastewater could also be used to supply make-up water for a power plant, which at the level of 1000 Mwe requires approximately 18 MGD of make-up water.

DEMAND FOR ELECTRICAL ENERGY

While the population of the United States has been growing at a relatively stable rate of about 25 million persons per decade over the past 20 to 30 years, the demand for electrical power has grown at a rate of about eight percent per year, doubling each decade. This phenomenon of disproportionate growth of population and electrical energy consumption is partially attributable to greater use of air conditioners, major appliances, television, commercial and street lighting, and a greater power consumption in the industrial sector. On a per capita basis, electrical power use is expected to increase over the next 40 years (8, 9). For Long Island, assuming the per capita electrical power demand will approximate the previous historical trend, and further assuming a relatively constant plant load factor (actual production of power divided by maximum possible production) of about 53 percent, it is possible to project the area power demand and the installed capacity needed to meet that demand. The Long Island Lighting Company's relatively low plant load factor (53 percent as compared to the national average of 64 percent) is attributed to the population variability caused by an influx of transients during the summer peak load period, and to the non-industrial character of the service area.

Electrical power requirements for Nassau and Suffolk Counties (10, 11) are shown in Table 4. The per capita use of electrical power in Nassau and Suffolk Counties was, in 1970, only about half the national average. Projected per capita power use, based on Edison Electric Institute and other projections for national averages, was accordingly reduced by 50 percent to correspond to the existing situation on Long Island. The plant factor was allowed to increase gradually from the 1970 value of 53 percent to 56 percent in 2020 in order to account for expected improvement in system efficiencies.

The 1970 generating capacity for Long Island was approximately 2330 megawatts. Subsequently, gas turbine capacity of 117 megawatts and 386 megawatts of fossil capacity were added in 1972. The Long Island Lighting Company has scheduled the Shoreham nuclear plant of 820 megawatts to enter service in 1976. These additions of 1323 megawatts will bring the system capacity to 3653 megawatts, substantially under the projected requirement for 1980 of 4740 megawatts. The difference will be accounted for by the construction of plants not yet announced or by the importation of power from interties with the New York Power Pool and the New England Power Exchange.

WASTEWATER

Latest data on wastewater flows in the Suffolk County area indicate that present requirements for wastewater treatment are on the order of 75 to 100 gallons per capita day (gpcd) (2). The lower figure represents current estimates of domestic and minor commercial consumption; the higher value includes an allowance for infiltration of groundwater into the sewer system. On this basis, if the County were to be entirely sewered, domestic wastewater flow would approximate 250 MGD by the year 2000, 300 MGD by the year 2020. In order to provide for treatment of wastewater resulting from increased industrial development such as that observed at present in western areas of Long Island (i.e., light "dry" industries such as electronics, metal fabrication, plastics, etc.), and for typical peak-to-average domestic consumption ratios on the order of 1.5, the wastewater treatment

TABLE 4

ELECTRICAL POWER DEMAND
FOR NASSAU AND SUFFOLK COUNTIES**

Year	Population	Per Capita Use Kw*	Total Use (Mw)	Plant Load Factor	Required Capacity Mwe*
1970	2,556,000*	0.43	1.10×10^{3}*	0.53*	2.33×10^{3}
1975	2,893,000	0.60	1.74×10^{3}	0.53	3.28×10^{3}
1980	3,221,000	0.78	2.51×10^{3}	0.53	4.74×10^{3}
1985	3,548,000	0.97	3.44×10^{3}	0.54	6.37×10^{3}
1990	3,933,000	1.20	4.72×10^{3}	0.54	8.74×10^{3}
1995	4,151,000	1.40	5.81×10^{3}	0.54	10.76×10^{3}
2000	4,336,000	1.60	6.94×10^{3}	0.55	12.62×10^{3}
2005	4,524,000	1.80	8.14×10^{3}	0.55	14.80×10^{3}
2010	4,718,000	2.00	9.44×10^{3}	0.55	17.16×10^{3}
2015	4,897,000	2.25	11.02×10^{3}	0.55	20.04×10^{3}
2020	5,071,000	2.50	12.68×10^{3}	0.56	22.64×10^{3}

*-Actual
**-Based on Long Island Lighting Company data, (10, 11)

capability must be at least twice the average domestic flow, or approximately 500 MGD by the year 2000, and 600 MGD by the year 2020.

The wastewater collection areas under consideration for supply correspond to sewerage Districts 11 and 13 described in Reference 2. These districts are approximately coincident with the surface drainage patterns.

District 11 encompasses an area of approximately 114 square miles and has a projected 1990 sewered population of 229,000 or an average density of about 3.1 persons per acre. This district is forecast to have a 1990 wastewater yield of approximately 40 MGD, or about 175 gpcd, including industrial, commercial, residential, and extraneous sources.

District 13 encompasses an area of approximately 27 square miles and has a projected 1990 sewered population of 85,000 or an average density of about 5 persons per acre. This district is expected to have a 1990 wastewater yield of approximately 9 MGD or about 107 gpcd including industrial, commercial, and residential sources. Tables 5 and 6 summarize the available data for the districts under consideration.

An integrated utility complex could therefore expect approximately 50 MGD from Districts 11 and 13 by 1990. Additional wastewater could be obtained from District 12, thus providing another 12 MGD, if the transportation cost were justified. Approximately 7 of a total of 17 miles of required trunk sewer could constitute a common system with the 11th District; such an arrangement would permit reaching 50 MGD prior to 1990.

Wastewater collection within the districts is planned to be primarily by gravity flow to one or two collection locations. The District 11 gravity collection location is planned for the Village of Riverhead, approximately at the junction of the Peconic River and Flanders Bay, and is about 7 miles from the proposed integrated facility. The static head between the collection location and the proposed plant site is estimated to be about 60 feet. Two gravity collection locations are planned for District 13; one at Mattituck and one at Greenport. The piping distance from Greenport to

TABLE 5

POPULATION AND WASTEWATER FLOW PROJECTIONS

District Number[1]	Year	Population[2]	Average Daily Flow (MGD)[2]	Peak Daily Flow (MGD)[3]	Minimum Daily Flow (MGD)[3]	Estimated Distance (miles)[4]	Static Head (feet)
11	1990	229,000	39.59	79.18	24.8	7	60
	2020	356,990	61.72	123.44	37.0	7	60
13	1990	84,000	8.85	17.70	5.3	15	60
	2020	218,300	18.98	37.96	11.4	15	60
12	1990	70,000	12.30	24.60	7.4	17	60
	2020	91,400	16.55	33.10	10.0	17	60

1 - District numbers correspond to the districts established by Bowe, Albertson, and Walsh in Reference 2.

2 - Based on Reference 2.

3 - Assumed peak-to-average flow of 2:1 and minimum-to-average flow of 0.6:1.

4 - Estimated piping distance from the point of district wastewater collection to the proposed treatment plant site.

TABLE 6

DISTRICT AREAS BY TOWNSHIP[1] (Square Miles)

District Number	Brook-haven	River-head	South-old	South-ampton	Total
11	42	50	--	22	114
13	--	--	27	--	7
12	--	--	--	25	25

1 - The sewage district boundaries are approximately the same as the surface water drainage divides, thus the surface drainage areas are roughly equivalent to the sewage district areas. The area figures are taken from "Comprehensive Public Water Supply Study, Suffolk County, New York, Vol. II, 1970" by Holzmacher, McLendon, and Murrell Consulting Engineers. (4).

Mattituck is approximately 12 miles and from Mattituck to the site is 3 miles, with a total increase in elevation of approximately 60 feet. Pumping and force main facilities would be required to transport wastewater to the treatment plant. Such facilities should have the capacity to deliver peak flows reliably as well as to maintain minimum velocities at low flow conditions to prevent the settling of solids.

SITE DESCRIPTION

The reference site for the integrated facility is a tract of land measuring approximately 4400 by 5200 feet (about 500 acres), located adjacent to the Southold Town Line at the northeastern corner of Riverhead Town, Suffolk County, New York. The site is at latitude 40° 59'30" North and at longitude 72° 36' West. The tract is bounded by Sound Avenue to the south, the Southold-Riverhead Town Line to the east, the Camp Carey access road to the west, and Long Island Sound to the north. The site encompasses almost all of the abandoned Camp Carey, and approximately 10 private residences. Major natural features of the site include Lily Pond, Hallocks Pond, and Jacobs Hill. With the exception of the range of sand hills directly adjacent to the shoreline, the site is relatively flat, averaging about 65 feet in height above sea level. A topographic map of the site is presented in Figure 5.

Since a nuclear plant is the choice for power generation, the site must be evaluated in terms of the Atomic Energy Commission criteria. These criteria require, among other things, a thorough demographic analysis of the region surrounding a potential site.

In the near vicinity of the site, population densities are quite low, ranging from about 0.5 to 3.0 persons per acre of land. Further west, toward the Nassau County line in the Town of Babylon, population densities reach 11.0 or more persons per acre, reflecting the spread of suburban communities eastward from New York City.

Within a 10-mile radius of the site, the largest community is Riverhead, with a 1970 population of 7585 permanent residents, and an anticipated growth to approximately 10,000 by 1980. This community is approximately six miles south-southwest of the site. The area within a 10-mile radius of the site is approximately 58 percent water.

Within a 20-mile radius of the site, the largest community is Yaphank in the Town of Brookhaven with a 1970 population of 8793 residents. The boundary of this community is approximately 20 miles west-southwest of the site. Neither Riverhead nor Yaphank are expected to have a 1980 population approaching 25,000 residents, the figure stipulated for a "Population Center" under the AEC criteria. Therefore, on Long Island, the distance to a Population Center is in excess of 20 miles, which represents a very conservative value. The approximate distance between the site and any point on the coast of Connecticut is also 20 miles. The distance to the nearest community of 25,000 or more residents is approximately 26 miles to East Haven, Connecticut with a 1970 population of 25,120 or New Haven, Connecticut with a population of 137,707 in 1970.

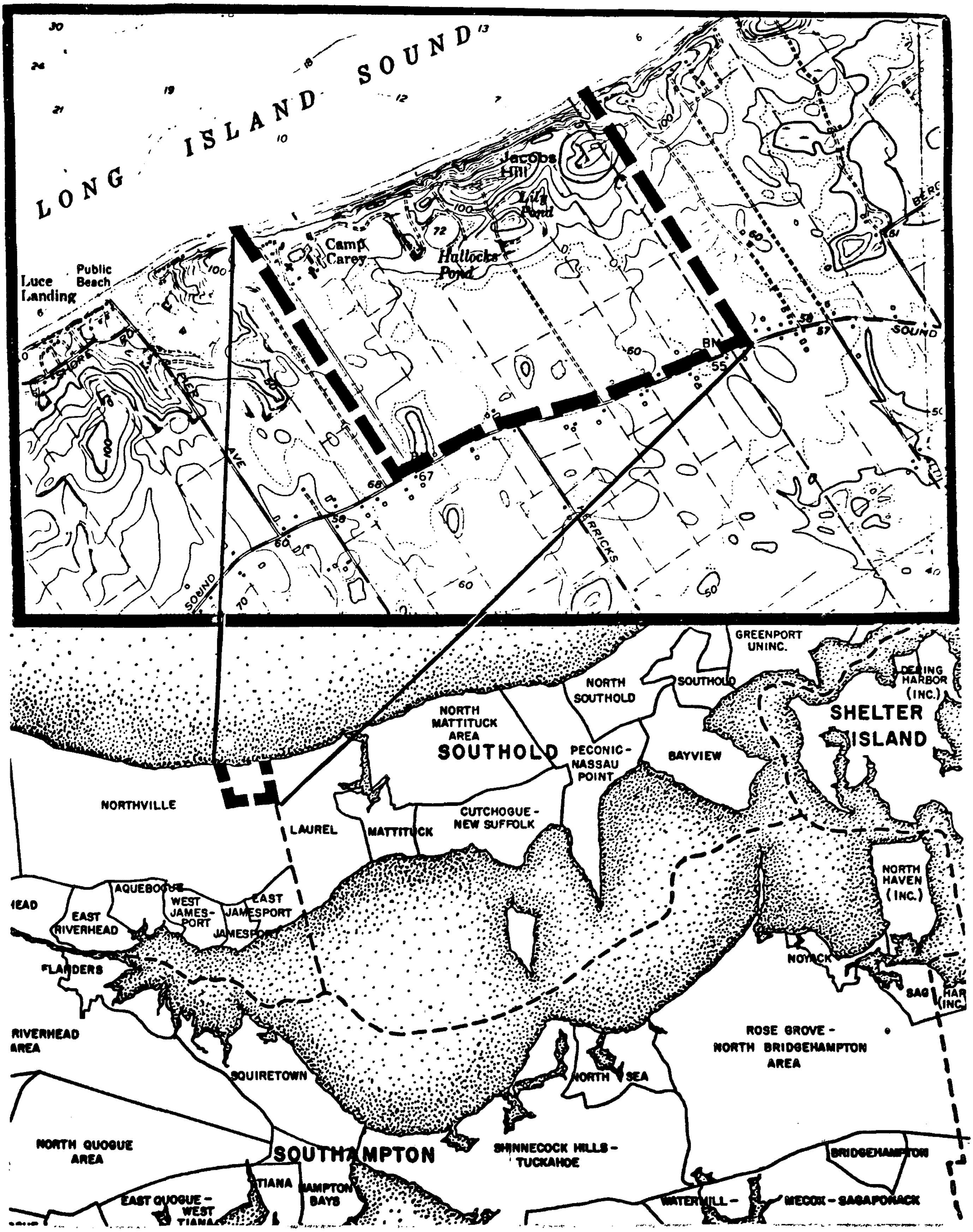


Figure 5. Reference Site Topography

Land use in Suffolk and Nassau Counties has been treated in detail by the Nassau-Suffolk Regional Planning Board (18). The information reported therein supports the conclusion that the site is surrounded by a sparsely populated area, as inferred from the population figures.

Land within a 10-mile radius of the site is within the Towns of Riverhead, Southold, and Southampton. Principal land use percentages for these towns and Suffolk County as a whole are shown in Table 7.

TABLE 7

PRINCIPAL LAND USE, PERCENTAGES

Categories of Use	Suffolk County	River-head	South-old	South-ampton
Residential	14	4	4	7
Commercial/ Industrial	2	1	1	2
Institutional	4	1	3	4
Agricultural	9	45	35	13
Recreational	7	8	7	6
Vacant (including waters)	54	23	42	62
Other	10	18	8	6
	100	100	100	100

From the physical and demographic data discussed previously, from a review of information previously compiled for preparation of a Preliminary Safety Analysis Report for a nuclear facility at the site (11), and from information presented in the LILCO Shoreham plant licensing application, the prospective multipurpose facility site may be characterized as follows:

1. The population density surrounding the site is low and is expected to remain so and the "Population Center" distance will be in excess of 20 miles; therefore, the

population distribution will not preclude site use.

2. Geological features of the area are relatively well documented. No faults are known to exist in the vicinity of the site and the nearest seismically active area is along the St. Lawrence Valley some 500 miles to the north.

3. Surface soil deposits at the site consist of unconsolidated sands; therefore, substantial attention to foundation design for site structures will be required.

4. Fresh groundwater lies at an elevation slightly above sea level at the site, or about 60 feet below the planned final grade. Drainage is directly northward toward Long Island Sound. Hydrologic conditions at the site are satisfactory.

5. Eastern Long Island is well-ventilated, with relatively high wind speeds. The metorological conditions are generally favorable for a nuclear site; however, consideration must be given in design for protection against hurricane force winds.

SECTION V

DESCRIPTION OF THE INTEGRATED FACILITY

The integrated plant design calls for the utilization of a 3400 Mwt light water reactor producing approximately 1000 Mwe energy for off-site utilization, providing process steam for a distillation plant capable of recovering approximately 47.5 MGD of high quality water, and providing all necessary electrical energy for the distillation and 50 MGD wastewater treatment plants. Steam from the final stage of the distillation unit will be used to elevate the temperature of the incoming raw wastewater and enhance the primary and secondary treatment processes. The secondary treated water will serve as feed for the distillation plant, which will recover 95 percent of the water that would otherwise be discharged to the sea.

The reference site selected for the design of the integrated complex encompasses over 500 acres and is located on the north shore of Long Island at the eastern end of the Town of Riverhead in Suffolk County.

Although the population density in the immediate area of the site and in most of eastern Suffolk County is relatively low, the rapid growth of population in Nassau County and in the western towns of Suffolk County will provide a demand for electrical power that appears to justify full utilization of economies of scale in sizing of the power generating facility. It is also clear that the future water requirements of the area can be augmented by the high quality water produced by the distillation plant. The factor which limits the water recovery capacity of the integrated facility is the supply of wastewater which can be economically conveyed to the facility for treatment. Waste Disposal Districts 11 and 13, comprising the Towns of Southold and Riverhead in the immediate vicinity of the reference site, the eastern sector of the Town of Brookhaven and the northern edge of the Town of Southampton, are projected to have a combined wastewater flow of approximately 50 MGD by the year 1990. Importation of wastewater from other districts in eastern Suffolk County could allow sizing of the wastewater treatment facility for a feed rate of 70 MGD or greater, achieving some further economies of scale. However, insufficient information relative to schedules for installation of sewers and the probable cost of conveyance systems for such importation is available at this time. The conceptual design of the integrated plant was therefore based on a feed rate of 50 MGD and is intended to allow expansion of the facility if required.

The nuclear power plant could utilize either a boiling water reactor (BWR) or a pressurized water reactor (PWR) as will be discussed in Section VI. A reboiler will be utilized to isolate the turbine feed steam, which potentially contains minor amounts of radioactivity, from the steam used in the distillation facility. Extraction of steam for this purpose will reduce the electrical output of the turbogenerator from the nominal 1100 Mwe associated with present generation nuclear plants to approximately 1000 Mwe. In all other respects, the nuclear power plant will be similar to those used for single purpose power generation. Radioactive waste disposal systems will be current state-of-the-art systems meeting all applicable regulations for environmental protection. Turbine condenser cooling will be provided by a once-through cooling system utilizing water from Long Island Sound and returning the water through an outfall structure designed to be totally compatible with environmental protection requirements. The nuclear steam supply system and the power generation and cooling system are described in Sections VI and VII, respectively, of this report.

Wastewater from Disposal Districts 11 and 13 (2) will be pumped to the site and will be treated in a modified activated sludge plant operating at an elevated temperature. Unit operations will consist of bar screening and pumping performed off-site, and thermal enhancement, grit removal, primary sedimentation, and activated sludge treatment (aeration and solids separation) performed on-site. No disinfection facilities, outfall sewer, or separate administration facilities are required. Primary and activated sludge will be combined, gravity thickened, digested, and dewatered on vacuum filters, then trucked to a landfill area.

The distillation plant, a 19-stage vertical tube evaporator (VTE) unit, will be operated on 285°F steam generated in a reboiler. The distillation plant will receive approximately 50 MGD of secondary treated wastewater, from which 47.5 MGD of product water will be produced. The difference, 2.5 MGD, will appear as evaporator bottoms with a total dissolved solids content of approximately half that of seawater, and will be mixed with approximately 1070 MGD of plant cooling water for discharge to Long Island Sound. The 50 MGD inflow of wastewater will be diluted with 1.44 MGD of distillation plant product steam, through injection by a barometric leg condenser, to provide thermal enhancement of the wastewater treatment process. Product water cooling required by the distillation plant will be provided by sharing the seawater intake and outfall system of the power generating facilities. Postdistillation carbon filtration, mineralization and chlorination will be included when the product water is to be provided for distribution. This post treatment will eliminate any volatile organics which may have carried through the distillation process and assure compliance of the product water with all water standards. The distillation plant system description is presented in Section IX of this report. An analysis of the economics of wastewater heating is presented in Appendix B.

A schematic flow diagram of the integrated plant is shown in Figure 6, and a preliminary site layout showing a possible arrangement of the nuclear power generating, wastewater treatment, and distillation facility on the proposed site is presented in Figure 7.

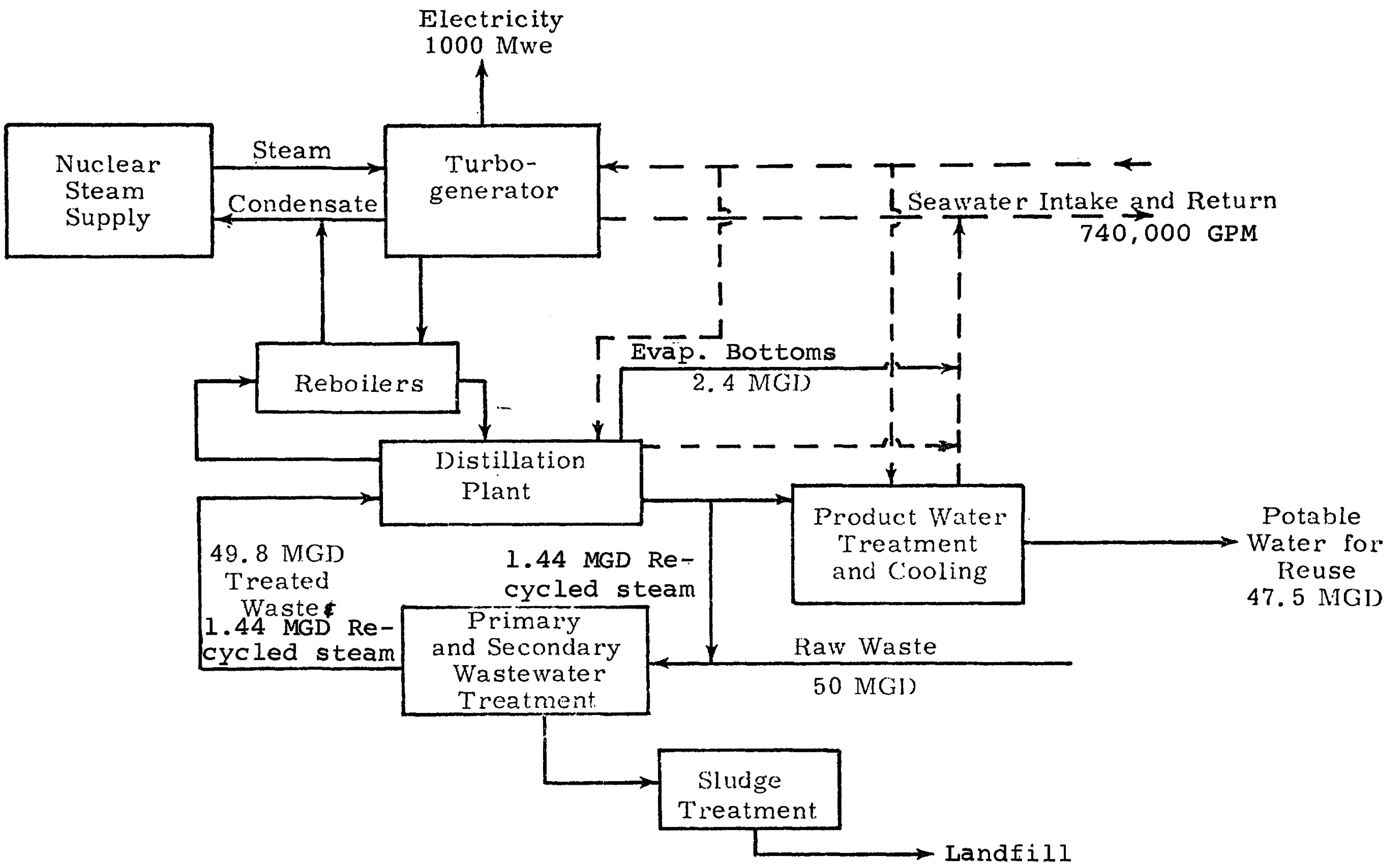


Figure 6. Schematic Diagram - Integrated Facility

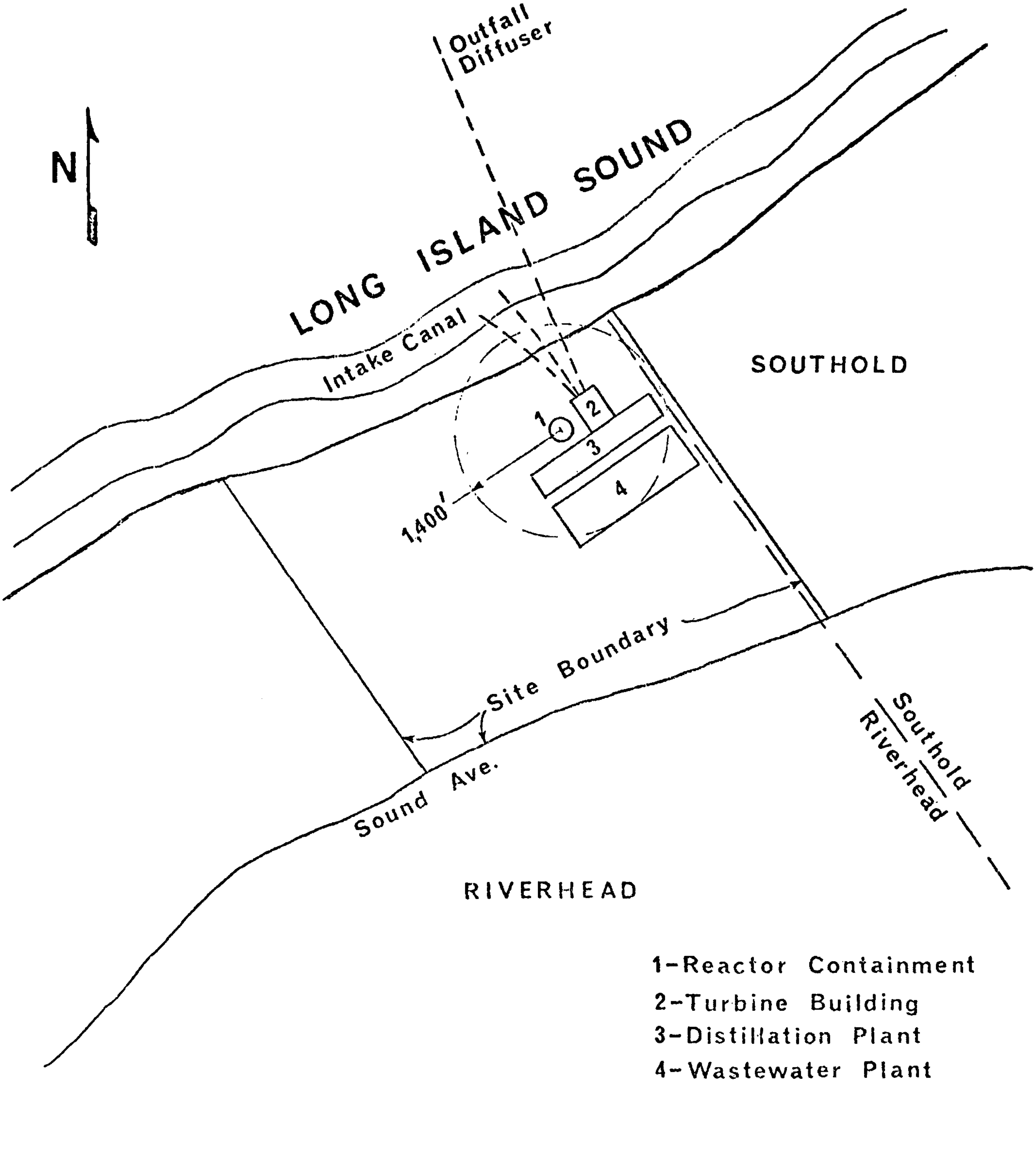


Figure 7. Site Plan - Integrated Facility

SECTION VI

NUCLEAR STEAM SUPPLY

Present and anticipated power demands in the region of the proposed integrated utility justify economy of scale in the selection of the nuclear power generating facility. On this basis, the nuclear steam supply system has been specified to be a 3400 Mwt reactor consistent with the capacity now being offered by reactor manufacturers. Further, for convenience in preparing this analysis, a light water reactor (LWR) was selected. This choice, however, does not constitute a requirement for a LWR in preference to a high temperature gas cooled reactor or other nuclear steam system. Data from two LWR vendors, General Electric and Westinghouse, were used to determine specific nuclear steam supply and turbogenerator characteristics and to develop estimates of capital and operational costs.

A 3400 Mwt nuclear system, when used for the single purpose of generating electricity, is capable of producing 1100 Mwe. The present application, however, calls for an integrated system, with steam being extracted for use in a distillation plant located on the same site as the reactor. This requirement reduces the net generating capacity of the system to approximately 1000 Mwe.

The choice of a LWR may be narrowed further to selecting either a pressurized water reactor (PWR) or boiling water reactor (BWR) for the proposed integrated facility. Initial consideration of both types of reactor indicates an advantage for the PWR for dual purpose use, in that steam generators are utilized to isolate the primary reactor coolant loop from the turbine steam system. Nevertheless, the nominal isolation in a PWR steam generator may be violated by small leaks, resulting in transfer of small quantities of radioactivity from the primary to the secondary system. With the conventional direct cycle BWR, steam generated in the nuclear reactor is fed directly to the turbine and therefore contains significant quantities of short-lived radioactivity.

Total isolation of wastewater reclamation process from the nuclear steam supply is considered to be essential, in order to achieve public acceptance. The approach to isolation selected for this study incorporates two high-reliability isolating steam generators, hereafter referred to as reboilers. The steam, extracted from the turbogenerator cycle of either the BWR or PWR, enters the reboilers where it generates the steam to be used in the distillation plant primary stage and in jet ejector service. The installed cost of these reboilers, including all associated instrumentation, is estimated to $2,500,000.

The main reboiler would operate on a nominal 10°F temperature differential, receiving 1.4 x 10^6 pounds per hour of 295°F process steam extracted from the turbine system and generating a like quantity of 285°F steam for use in the distillation plant. A second reboiler operating on 385°F steam would be used to generate approximately 10,000 pounds per hour of 100 psi steam for use in the distillation plant ejectors.

These reboilers and the provisions for interstage extraction of steam for their operation constitute the only major non-standard features of the nuclear steam supply and power generating systems. Radioactive waste monitoring, treatment, and disposal features of the plant will be specified and designed to meet all requirements of the USAEC and other regulatory agencies, as applicable to the site. Similarly, design features for environmental protection will be incorporated in accordance with standard nuclear design practice and the special requirements of the specific site location.

Table 8 summarizes a current estimate of nuclear plant costs (15). The tabulated costs are separated into components associated with the nuclear steam source and the turbogenerator. A 15 percent fixed charge rate was assumed to be appropriate for a New York State private utility and all capital and construction costs were based on 1972 dollars and using an Engineering News Record Construction Cost Index (ENR) 1690.

The levelized fuel cost for the nuclear power plant has been selected at 20¢/10^6 Btu on the basis of the values reported by Westinghouse Electric Corporation (16). Other sources

TABLE 8

NUCLEAR PLANT INITIAL INVESTMENT COST ESTIMATE
FOR 1972 OPERATIONS
1100 Mwe SIZE*

	Steam Generator $/Kwe	Turbo-Generator $/Kwe
Unit Cost	40	31
Construction Materials and Equipment	27	18
Construction Labor	33	23
Additional costs for 1972**	22	--
Professional Service	14	11
Other Indirect Costs	12	9
	148	92
Escalation During Construction	0	0
Interest During Construction	27	17
Total	175	109
Nuclear Plant Cost	$284/Kwe	

*-Based on Reference 15.

**-Includes regulation and safety, near-zero radiation release, additional quality control requirements, and aesthetics.

(15, 17) generally support that cost, citing a range from 17.9¢/10^6 Btu to 22.1¢/10^6 Btu, with an average of 20.6¢/10^6 Btu.

The cost of prime steam includes the cost of the nuclear steam generator, fuel costs, and the portion of the operating and maintenance costs attributed to steam production.

Annual Cost of Steam Production

Steam Generator \$175/Kwe (1.1×10^6 Kwe) (15% Fixed Charge Rate)	=	$\$28.9 \times 10^6$/yr
Operation & Maintenance @ 0.695 mills/Kw-hr	=	5.3×10^6/yr
Fuel @ 20¢/10^6 Btu	=	16.2×10^6/yr
		$\$50.4 \times 10^6$/yr

Prime Steam Cost (519°F)

Total Annual Steam Production = 8.1×10^{13} Btu

$$\text{Cost} = \frac{\$50.4 \times 10^6 \text{ per year}}{8.1 \times 10^{13} \text{ Btu}} = 62.3¢/10^6 \text{ Btu}$$

The cost of process steam for the distillation plant may be estimated on the basis of relative energy utilization. The base cost of steam is 62.3¢/10^6 Btu. This is the cost of steam assuming that no energy is utilized for the production of electricity. However, the available energy of prime steam is much higher than that required for the distillation plant. Thus, useful work can be performed by the steam before it is extracted for process purposes. Using the Carnot efficiency approach the amount of energy extracted from the steam can be approximated to vary linearly with the extraction end point. Thus, the energy of 519°F steam will cost 62.3¢/10^6 Btu and 101°F exhaust steam (the turbine terminal temperature) will cost nothing. The cost of steam of intermediate temperatures may be approximated by a straight line function between these two extremes. The equation of this line is:

$$C_{steam} = \frac{62.3¢}{10^6 \text{ Btu}} \times \frac{T_s - 101}{519 - 101}$$

Thus, the cost of 295°F steam to the main reboiler is 29¢/10^6 Btu, and the cost of the small additional quantity of 358°F steam required for the high pressure reboiler is approximately 38¢/10^6 Btu.

SECTION VII

POWER GENERATION AND COOLING SYSTEM DESCRIPTION

The 3400 Mwt nuclear reactor, chosen as the energy source for the integrated facility, is typical of the light water reactors currently being purchased for commercial operation in this country, takes advantage of the economies of scale and reduces the operating cost per unit of electrical output. The relative ease of transmitting electricity, as compared to water, permits the electrical plant to serve a significantly larger service area and population than will the wastewater and distillation plants.

The nuclear steam supply would permit production of 1100 Mwe if it were used in a single purpose facility. The multipurpose facility described in this study will have a net electrical output of approximately 1000 Mwe with the distillation plant and wastewater treatment plant in full operation. The turbogenerator to be used with the plant will be a conventional 1000 Mwe type, modified for interstage extraction of process steam at temperatures of 295°F and 385°F. This steam is utilized in isolating reboilers to provide both process steam to the distillation plant primary stage and steam for operation of the ejector system throughout the distillation plant. With the exception of the provision for extraction of steam, the turbine is fully condensing, utilizing once-through seawater as a condensing fluid, as discussed below, with terminal steam conditions of 101°F and 2" Hg pressure. An operating efficiency for the turbogenerator of approximately 35 percent may be anticipated.

A turbogenerator of the required characteristics can be supplied by any of a number of manufacturers. The specific design characteristics would depend on the manufacturer as well as the nuclear steam supply system selected.

Recent estimates (15) indicate a total cost of approximately $109,000,000 for the turbogenerator; i.e., approximately $109 per installed kilowatt of electrical power. This cost includes approximately $30 per kilowatt as the basic unit cost; approximately $40 per kilowatt for construction labor, materials, and equipment; approximately \$20 per kilowatt for engineering professional services, and other indirect costs; and up to \$20 per kilowatt for interest during construction (all cost estimates are expressed in 1972 dollars using an ENR index of 1690).

Costs of the nuclear portion of the power generating station, described in Section VI are estimated to be \$175 per equivalent electrical kilowatt; i.e., since the nuclear station is sized for the equivalent of an 1100 Mwe generating capacity, the total estimated cost is approximately \$192,500,000. Combining the steam supply cost with that of the turbogenerator and auxiliary equipment at an annual fixed charge rate of 15 percent, fuel cost at 20¢/10^6 Btu, and operational and maintenance cost at 0.93 mills/Kw-hr, leads to a total bus bar electric cost for the plant of approximately 9.1 mills/Kw-hr, as follows:

Annual Fixed Costs (7000 hr/yr Operation)	6.1 mills/kw-hr
Fuel (10,390 Btu/kw-hr)	2.1 " "
Operation and Maintenance	0.9 " "
	9.1 mills/kw-hr

The general energy balance for the integrated facility is shown in Figure 8. As indicated, the total energy production is approximately 11.4 x 10^9 Btu/hr. Of this, 3.4 x 10^9 Btu/hr is transmitted off-site in the form of electrical energy. The remaining energy, 8.0 x 10^9 Btu/hr, although utilized in the wastewater treatment and distillation processes, (e.g., distillation of water and operation of pumps and other electrical equipment) is not consumed and must, therefore, eventually leave the site in the form of heat. An insignificant quantity of heat, approximately 3.9 x 10^4 Btu/hr leaves the wastewater process with the sludge and is eventually lost to the atmosphere. It should be noted, however, that approximately one-third of this heat, 1.3 x 10^4 Btu/hr. is produced by burning of methane gas given off in the sludge treatment process and used to increase the reaction rate in the sludge digesters. Approximately 2 x 10^8 Btu/hr. of heat is lost to the atmosphere from the distillation and wastewater treatment process by radiation and convection from the process equipment.

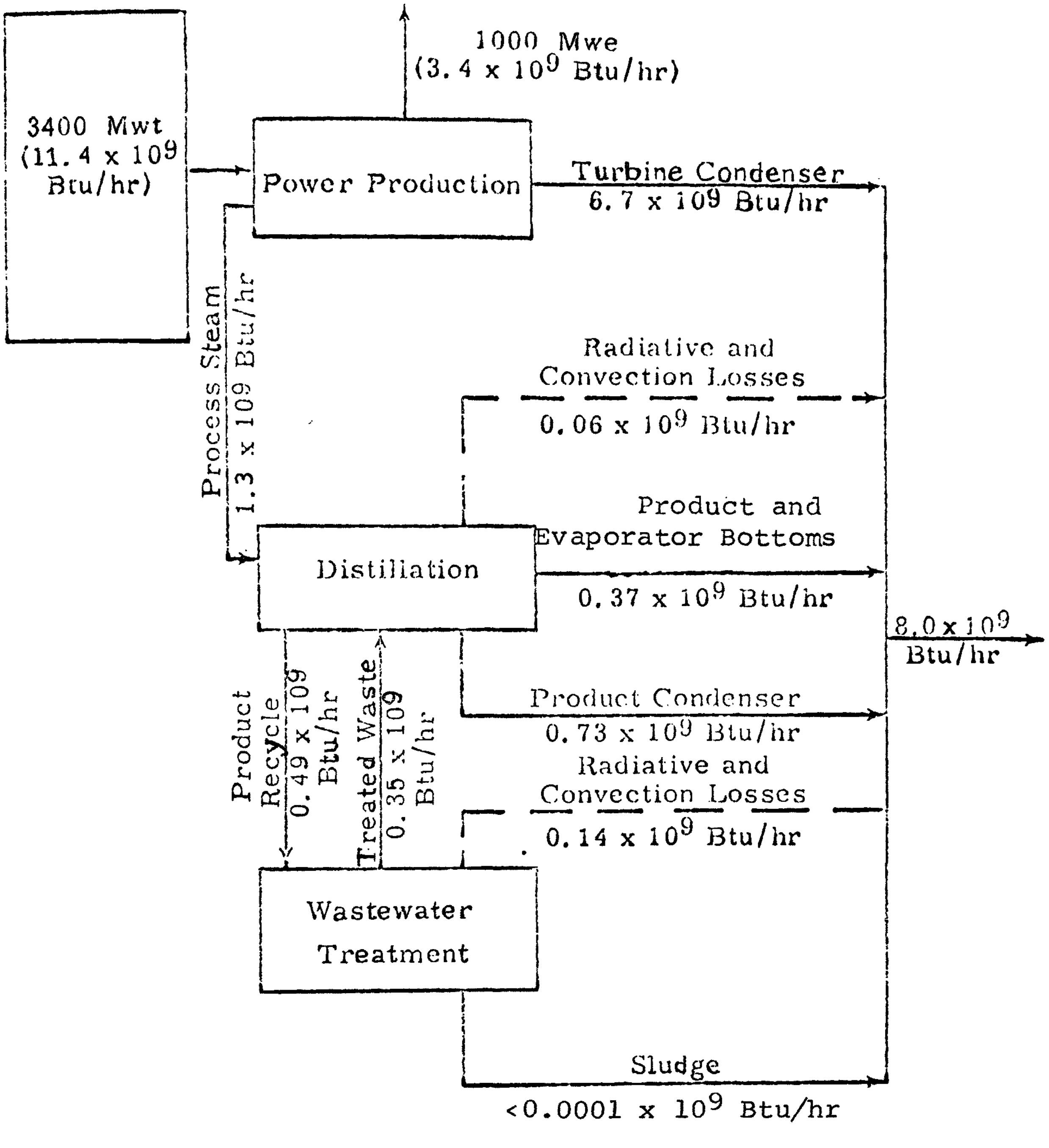


Figure 8. Energy Balance - Integrated Facility

Evaporator bottoms and product water leave the site at temperatures of 35°F and 20°F, respectively, above the temperature of the incoming raw wastewater, carrying with them a total of approximately 3.7×10^8 Btu/hr. The remaining quantity of heat, approximately 7.4×10^9 Btu/hr, must be removed and discharged to the environment through use of a cooling system. The primary source of this waste heat is the power station turbine condenser discharging 6.7×10^9 Btu/hr. The remainder of the heat, 7.3×10^8 Btu/hr, must be removed from the distillation plant product condenser and product water cooling system.

The simplest and most direct approach to condenser cooling is to utilize water from Long Island Sound on a once-through basis. A flow of approximately 740,000 gpm will be required if the temperature rise across the condenser is to be limited to 20°F, a value which will be near optimum from engineering and economic considerations. If Long Island Sound water is to be used for cooling, a large and enviromentally compatible intake structure must be provided and provision must be made for an outfall system which will dissipate the heat at an approximate depth and over an adequate area to minimize environmental impact and meet applicable regulations.

From the Shoreham Nuclear Station now under construction, the Long Island Lighting Company has designed an outfall system to conform to the regulations governing thermal discharges. A single pipe will extend from the plant for a distance of 1600 feet along the floor of the Sound. From that point, the outfall line will be raised three feet from the floor of the Sound. The pipe will run an additional 2200 feet in a northerly direction into the Sound in deep water. Outlet ports will be spaced every 60 feet from the elevation point and will be alternated on either side of the pipe to achieve maximum dispersion of the thermal effluent.

The evaluation performed for the Long Island Lighting Company Shoreham Plant suggests that once-through cooling, with discharge of waste heat to Long Island Sound, would constitute an environmentally acceptable approach for the integrated facility under evaluation. The environmental, engineering, and economic feasibilities of alternatives to this approach have also been evaluated.

The alternative to the discharge of heat to Long Island Sound is discharge to the atmosphere, either by evaporative or non-evaporative cooling towers. Dry (non-evaporative) cooling towers cost from $30 to $45 per installed kilowatt and have never been utilized on a plant of the size proposed here. Moreover, dry cooling requires an increase in terminal turbine steam condensation temperature, resulting in a loss in turbine efficiency and a concomitant economic penalty (18). For these reasons, the use of non-evaporative cooling towers was not considered further.

Evaporative cooling systems may be grouped into three general categories: natural draft towers; mechanical draft towers, and spray ponds. Recirculating cooling ponds other than spray are discussed below. Cooling ponds transfer heat to the atmosphere both by evaporation and by radiation and convection, with evaporation typically resulting in approximately one half of the heat transfer. This fraction, and the successful operation of these systems, is highly dependent on local meteorological conditions.

Evaporative cooling systems are characterized by transfer of heat from water to air in the form of latent heat of evaporation. The disposal of 7.4×10^9 Btu/hr requires, therefore, the evaporation of approximately 13,000 gpm of water for a total evaporative loss of approximately 18 MGD. If fresh water supplies, groundwater, or other water of potable quality were to be utilized, this arrangement would represent a significant economic loss. Utilization of seawater from Long Island Sound would appear to be more economically attractive; however, the extent and effect of salt drift from the evaporative cooling systems is a factor which must be evaluated (20).

One alternative which was given major consideration during this study was the use of partially treated wastewater for condenser cooling purposes, transferring heat to the atmosphere utilizing a cooling pond with floating spray modules. This approach also represents an integration of wastewater treatment, power production and water recovery facilities. In addition to providing a means of heat dissipation, aeration serves to further the wastewater treatment process. Partial treatment of the wastewater appears to be necessary, at the present time, because of the uncertainty in values for the corrosion and fouling.

Another advantage initially credited to this approach was the elimination of the requirement for a specific ammonia removal step in the wastewater handling. Subsequent analysis, however, discussed in Section IX, showed that ammonia could be removed directly in the distillation process and that treatment of the wastewater for this specific purpose was not required.

A disadvantage of the spray pond system is the space requirement of approximately 30 acres of active spray pond and additional land for approach and return channels. Considering that the study site contains over 500 acres, this space utilization appears to be acceptable. A second and more significant disadvantage of this system as an alternative evaporative cooling system is the unavoidable loss of approximately 36 percent of the wastewater collected and treated which, with further treatment and distillation, could be recycled.

The economics of closed cycle cooling have been studied extensively by Hittman Associates (18) and others (21). Application of these data to the integrated facility would increase the electrical requirements for a closed cycle spray module system by the equivalent of $1400 a day. This cost, however, is secondary to the loss in revenue resulting from an estimated 1.8 percent reduction in power generation attributed to the increase in turbine heat rate from a value of 10,390 Btu/Kw-hr with once-through cooling to an estimated 10,575 Btu/Kw-hr on the closed cycle. Based on the bus bar cost of 9.1 mills/Kw-hr, this production loss is equivalent in value to almost $4000 per day.

Theoretically, the calculated spray module cost of $5,300,000 could be more than offset by elimination of much of the $12,200,000 intake/outfall structure required for the plant. Complete elimination of the intake/outfall structure is not practical, however, since some provision for discharge of secondary treated waste during periods of distillation plant outage and of distillation plant evaporator bottoms during normal operation is required. Alternatively, the 2.4 MGD of 15,000 ppm concentrated evaporator bottoms could be further concentrated, dried, and disposed of by incineration or landfill, if desired.

The use of a fresh water and cooling water storage reservoir as the primary condenser cooling system was also considered.

This reservoir would be supplied not only by product water from the distillation plant, but also by local surface and groundwater supplies. The evaporative loss from such a reservoir, assuming recirculation through the plant condenser, would be approximately 10 MGD. This loss would be composed entirely of high quality water which could otherwise be returned directly to the Suffolk County water supplies. The reservoir would require approximately 1800 acres, or nearly three square miles. Topographic factors in the vicinity of the site are not conducive to installation of such a reservoir. On the basis of these factors, this alternative was not considered further in this study.

In summary, the power generation system will consist of a multistage condensing turbine approximating conventional nuclear design and with an electrical generating capacity of 1000 Mwe minimum. Provisions will be made for interstage extraction of 295^{o}F saturated steam at rates up to 1.4 million pounds per hour, and 385^{o}F steam at a much lower rate of 10,000 pounds per hour for use in the distillation process. Based on a nominal operational schedule of 7000 hours per year, the plant will produce 7×10^9 kilowatt hours of electricity per year for off-site use, at a bus bar cost of 9.1 mills/Kw-hr. Waste heat from the turbine condenser will be discharged to Long Island Sound using a once-through cooling system. With a nominal seawater inlet temperature of 71^{o}F, a 20 degree temperature rise across the condenser, and a 10 degree approach, the turbine will be operated under condensing conditions of 101^{o}F and 2" Hg abs, with an anticipated heat rate of less than 10,390 Btu/Kw-hr. Protection of the aquatic environment from the temperature discharge will be provided by limiting the temperature rise across the condenser and providing a suitable outfall structure. Total cooling water utilization for the power and distillation plants will be approximately 740,000 gpm.

SECTION VIII

WASTEWATER TREATMENT

The proposed wastewater treatment plant is designed for an average flow of 50 MGD, with a peak hydraulic capacity of 75 MGD. Minimum flows are expected to be on the order of 30 MGD. Heat is to be added by means of a barometric condenser at the inlet of the plant.

Influent BOD and suspended solids are estimated to be 200 mg/l and 235 mg/l respectively, and effluent BOD and suspended solids are projected to be 20 mg/l each. The plant is designed to remove 90 percent of the incoming BOD or 75,000 pounds per day and 91.5 percent of the suspended solids or 89,500 pounds per day. Sludge handling and disposal facilities are designed to accommodate 55 tons of solids per day.

The process flow sheet for the proposed plant is shown on Figure 9. Because the plant is located on the north shore of Long Island, adjacent to the Southold town line at an elevation of approximately 60 feet above sea level, all wastewater will be pumped to the plant. Pumping stations will be located at low points in Districts #11 and #13 (2), and will be provided with facilities to screen the raw wastewater. The first unit operation at the plant site will be thermal enhancement by means of barometric condensers. Steam from the 19th stage of the distillation plant will be condensed to raise the temperature of the raw waste to 93°F in order to assure a temperature of 86°F (30°C) in the aeration basins of the activated sludge process.

The grit chamber is designed to remove grit from the thermally enriched wastewater. The hydraulic capacity of the grit chamber was increased by approximately three percent to allow for the increase in flow due to condensation, but the size of the unit has been reduced by 30 percent to reflect the improved performance due to the elevated temperature.

Primary sedimentation tanks are designed to remove the settleable solids from the degritted wastewater, including the condensate. The sedimentation tanks were designed on the

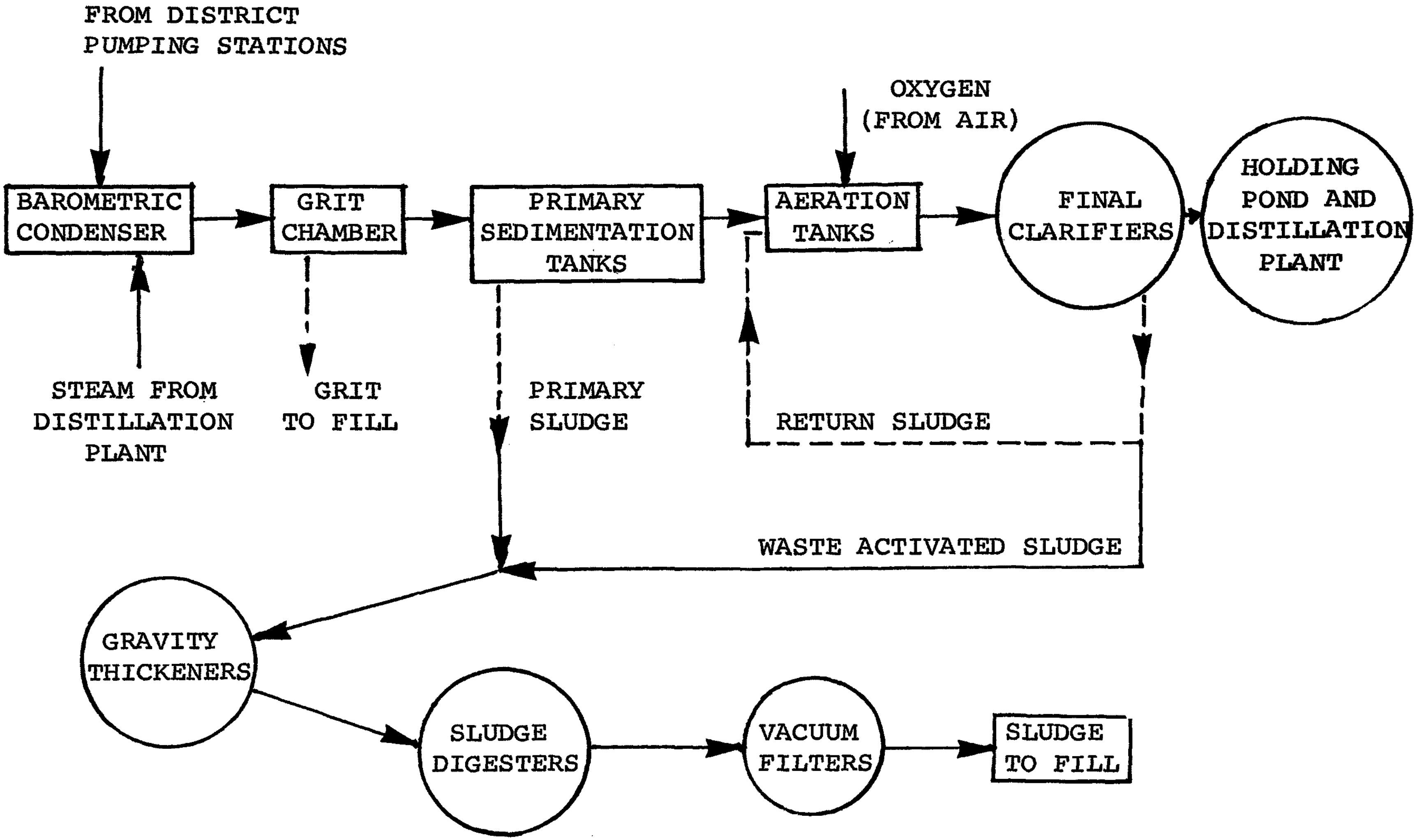


Figure 9. WASTEWATER TREATMENT PLANT PROCESS FLOW SHEET

basis of an elevated temperature of 92°F and are approximately 75 percent of the size of conventional primary sedimentation tanks designed for operation at ambient temperature.

The activated sludge units have been designed to operate on heated waste at approximately 86°F (30°C) and reflect a reduction in aeration tankage of approximately 30 percent. The required aeration facilities are comparable in size to conventional equipment because the increase in oxygen transfer rate is offset by the decrease in oxygen saturation at elevated temperatures.

Final clarifiers have been reduced 20 percent in size, as a result of better liquid-solids separation at the elevated temperatures of 85°F.

The discharge from the final clarifiers passes directly without disinfection to a buffer reservoir ahead of the distillation plant. No separate outfall is required because the entire effluent is used in the distillation plant. In the event that the distillation plant is shut down, the effluent is chlorinated with the disinfection facilities of the distillation plant, using the buffer reservoir as a chlorine contact tank. The wastewater is then discharged with the cooling water from the power plant condenser.

Primary sludge and waste activated sludge are combined and thickened in gravity thickeners which have been reduced in size, as a result of the 85°F elevated temperatures of the wastewater, to approximately 80 percent of the capacity of conventional gravity thickeners operating on unheated sludge. Thickened sludge passes to anaerobic sludge digesters where approximately 70 percent of the volatile solids are destroyed, yielding sludge gas (methane and carbon dioxide) which is used to heat the sludge digesters to 95°F and can be used to heat the buildings and the air supply for the ammonia stripping process.

Digested sludge is dewatered on vacuum filters. The sludge cake, i.e., the product from the vacuum filters, is trucked to landfill. Selection of landfill as the final disposal method for the dried sludge is optional and does not influence the objectives of this study. Alternative means of final sludge disposal include incineration, disposal at sea, and wet air oxidation.

The barometric leg condenser was selected for incorporation into the integrated project because under the constraint of current technology it is the most economical method of heating wastewater. The proposed treatment process will utilize heat from the distillation plant and will effectively remove contaminants from the wastewater. The wastewater treatment plant effluent is projected to contain approximately 20 mg/l of NO_3 and 5 to 10 mg/l of NH_3. The effluent turbidity is estimated to be 5 to 10 Jackson Turbidity Units (JTU). The estimated capital costs for the wastewater treatment plant in the integrated complex are compiled in Table 9, which summarizes the integrated system with treatment plant heat supplied by barometric condensers.

Capital costs reflect the construction of a plant to treat waste at an elevated temperature and do not include disinfection facilities, outfall sewer, and administration facilities which would normally be required for a conventional activated sludge plant. Disinfection is not required insofar as the subsequent process is one of distillation. The outfall has been eliminated because the entire output of the wastewater treatment plant will pass through the distillation plant, or through the outfall of the power plant cooling water system in the event of a shut down of the distillation plant. The administration facilities of the integrated facility are considered to be adequate to service the needs of the wastewater treatment plant.

TABLE 9

50 MGD WASTEWATER TREATMENT PLANT COST ESTIMATE
INTEGRATED SYSTEM, PLANT HEATED BY BAROMETRIC CONDENSERS*
ENR 1690

Raw Waste Pumping Stations	$ 2,415,000
Barometric Condenser	Included in distillation plant cost-see Chapter IX
Grit Chambers	580,000
Primary Settling Tanks	810,000
Primary Sludge Pump Station	130,000
Aeration Basins	2,240,000
Aeration Equipment	1,280,000
Final Clarifiers	1,055,000
Return Sludge System	395,000
Gravity Thickeners	340,000
Anaerobic Sludge Digesters	2,415,000
Vacuum Filter Facilities	1,790,000
Maintenance Facilities	160,000
Yard Piping	2,000,000
Subtotal	$15,610,000
Engineering and Contingencies @ 30%	4,685,000
Total	$20,295,000

*-Inlet to grit chamber, 93°F; primary settling tanks, 92°F; aeration basins, 86°F (30°C), plant outlet, 85°F.

SECTION IX

DISTILLATION PLANT

SYSTEM DESCRIPTION

The distillation process employed in the integrated facility is a falling-film, multiple-effect, vertical tube evaporative (VTE) process with 19 heat recovery effects. Accumulating condensate and wastewater feed are flashed separately in a single flashing stage for each effect. The plant has a performance ratio of 14.5 pounds of product water for each 1000 Btu of input steam. The maximum distillation temperature of the first stage concentrate is approximately 275^{o}F. Other temperatures and flow rates are shown on the Process Flow Diagram, Figure 10.

Input to the distillation plant is approximately 50 MGD of treated wastewater and 1.4 MGD of distillation plant exhaust steam previously condensed in the wastewater. Similarly, the design output of 47.5 MGD does not include the 1.4 MGD of steam withdrawn from the final stage to heat the wastewater. The difference between the input and output, approximately 2.5 MGD, consists of sterile evaporator concentrates and system losses. The evaporator concentrates, which contain dissolved solids at approximately 15,000 ppm concentration are discharged through the turbogenerator cooling water outfall to the waters of Long Island Sound, which typically contain 33,000 ppm of dissolved solids. Steam condensation and product water cooling required at the distillation plant will be provided by once-through cooling with seawater utilizing the same intake and outfall facilities as the power generating system. After distillation, the product water is subjected to treatment such as carbon filtration and chlorination, depending upon the intended use.

The secondary treated wastewater is pumped from the final clarifier, located in the wastewater treatment facility, to a decarbonation basin. Before entering this basin, sulfuric acid is added to neutralize its alkalinity. In this neutralization reaction, the bicarbonates are decomposed and free

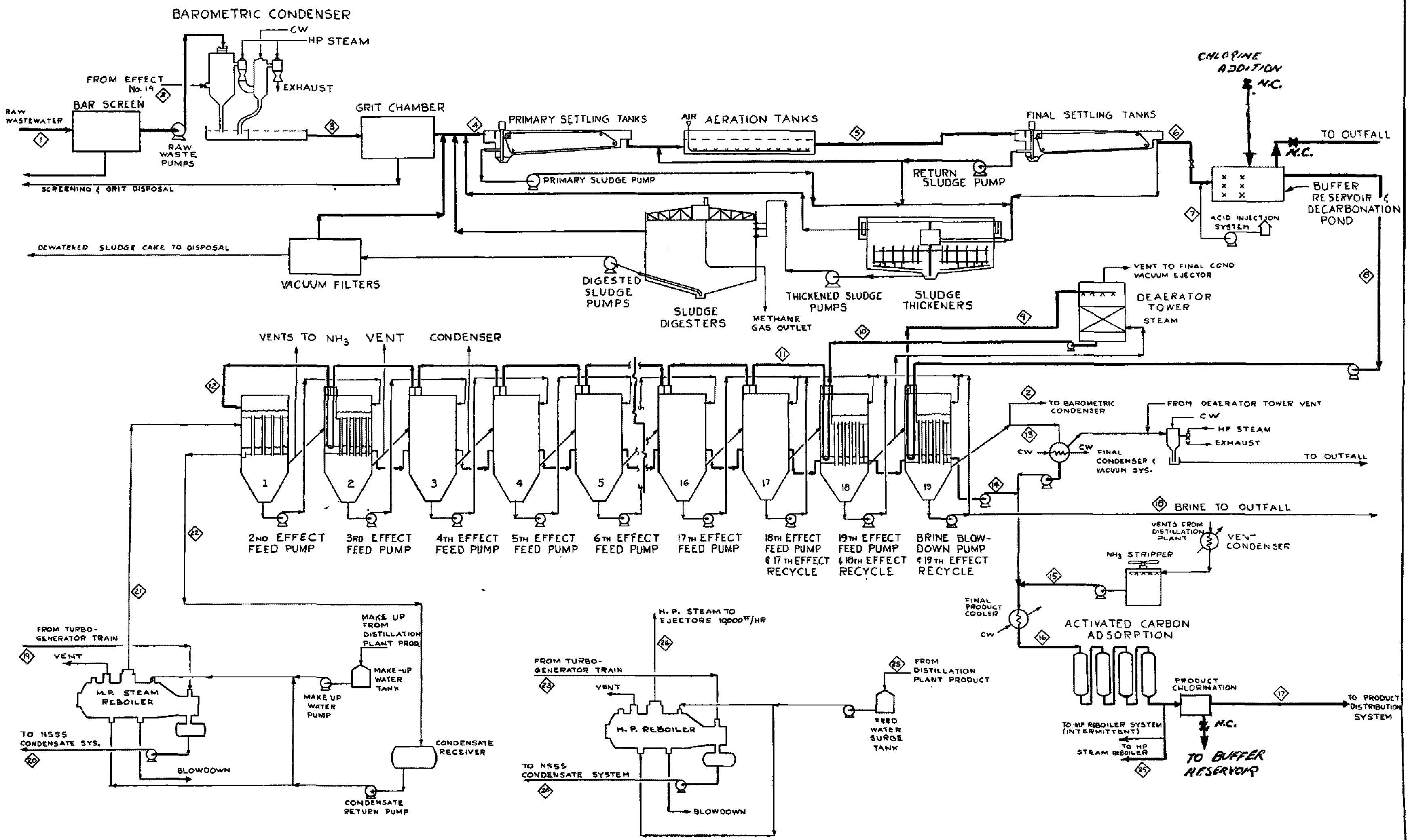


Figure 10. INTEGRATED FACILITY OVERALL HEAT & MATERIAL BALANCE DIAGRAM

LINE NUMBER	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
FLOW, MGD	50.0		51.44	55.15	67.24	51.44		51.44	51.44	51.44	51.44	51.44		45.47		47.5	47.5	2.9								
FLOW, LB/HR.		499,425					2,140						703,575						1.395×10^6		1.384×10^6		13,150			10,000
FLOW, GPM															AC					2790		2770		27	20	
TEMPERATURE, °F	65	100	93	92	86	85	AMB.	84	94	94	104	275	100	110	150	85	85	100	295	290	285	280	358	353	85	350
PRESSURE, PSIA																			62.1	58.0	53.2	50.0	150	148		100
PRESSURE, INCHES OF HG ABS		2.0											2.0													

carbon dioxide is evolved. The decarbonation basin consists of a set of stair-like steps leading down to a retention pond. The water is introduced at the top step and cascades down the steps into the retention reservoir, from whose opposite end the water is withdrawn. The fall-splash process at each step continuously renews the liquid surface to effect good gas release. The design for the proposed plant consists of a weir, five steps (each 3 feet tread, 3 feet rise, and 40 feet wide), and a retention tank (200 feet long, 40 feet wide, and 4 feet deep). Total retention time is approximately seven minutes. Head loss is expected to be approximately 8.5 psi.

The decarbonation basin is to be constructed as part of the entrance to the buffer reservoir. This buffer reservoir is expected to hold approximately five hours of normal flow, or about 10 million gallons. The approximate size is 200 feet wide, 670 feet long, and 10 feet deep. From the decarbonation basin and buffer reservoir the decarbonated feed is introduced to the last effect preheater section, where its temperature is raised to approximately 94°F, prior to entering the deaerator tower. The reasons for deaerating to low levels (10 ppb of oxygen and 1 ppm of carbon dioxide) are (1) to minimize corrosion of the evaporator internals; (2) to minimize condenser gas fouling; and (3) to prevent formation of carbonate scale.

The deaerator proposed for this plant is a packed tower utilizing stripping steam for gas release. The equipment consists of a vertical cylindrical column containing a bed of packing which breaks up the water and maximizes the exposed surface.

In the deaeration process, the water is spread and sprayed over the top of the packing bed. As the water trickles down through the packing, stripping steam is supplied at the bottom and travels upward through the packing. The gases are separated from the water and are carried away by the stripping steam in accordance with Henry's Law for concentration of a dissolved gas. If the steam supply is pure, the concentration of the dissolved gases in the water can be made to approach zero. An average liquid loading rate of 20,000 lbs/hr-ft^2 is considered practical. The flow rate for the stripping steam is a function of the degree of deaeration required, the amount of steam required to minimize channeling, and the characteristics of the particular type of packing. The deaeration capacity required for this plant is provided by two 25-foot diameter by 40-foot high towers in parallel. (22, 23).

To the deaerated feed is added an antifoaming agent to improve heat transfer. The feed is then pumped through successive preheaters positioned in each effect arriving at Effect No. 1 at a temperature of 275°F where it enters the first bundle of vertical falling-film tubes, and proceeds through the 19 stages of the VTE plant. The VTE plant receives its heating steam from an intermediate steam reboiler, and all of the condensate from the heating steam is returned to this reboiler. This steam (approximately 1.4×10^6 lbs/hr of reboiled steam) at 285°F and 53.2 psia (saturated) is admitted to the shell-side of the Effect No.1 falling-film bundle. The steam condenses on the tubes, releasing its heat of condensation to the preheated, decarbonated and deaerated feedwater which is falling as a film on the interior surface of the tubes. This causes the feedwater to boil violently, so that approximately five percent is vaporized as it falls through the vertical tubes. The mixture of wastewater and newly formed steam flows from the bottom of the heater tube bundle into the lower section of the Effect No. 1 evaporator where the wastewater disengages from the steam and collects in the sump. The slightly concentrated feed is then pumped to the top of Effect No. 2 while the steam is directed to the shell-side of the Effect No. 2 vertical tube bundle and feed preheater tube bundle.

Knitted wire mesh entrainment separators are provided to remove entrained droplets of liquid from the steam flow. In Effect No. 2, the vapor condensing on the outside of the vertical tubes performs the same function as the heating steam did in Effect No. 1, boiling more feed on the inside of the tubes and producing additional steam to be passed on to Effect No. 3. These essential processes are repeated in each of nineteen effects through the plant, with each effect operating at a progressively lower temperature and pressure. In all effects (except the first), the falling-film tubes are supplied by a pump which withdraws the concentrated feed from the sump and lifts it to the top of the vertical bundle. Special distributors (slotted circular weirs) distribute the concentrated liquid to the tubes through which it returns, by gravity, to the sump.

Ammonia removal is accomplished primarily through venting the initial three or four stages. Continuous venting of all effects is necessary to remove noncondensable gases. Effects operating above atmospheric pressure are vented to a condenser where the vapors are condensed (except for noncondensables which are released to the atmosphere) and fed to an ammonia stripper where any remaining ammonia is removed. The clean liquid that leaves the stripper is combined with the other product water streams. The condensed vent vapor flow is approximately 50 to 100 gpm and is at a temperature of 150°F when combined with the other product water upstream of the product water cooler.

Approximately 58 percent of the 100°F steam that is formed in the last effect (No. 19) is condensed in a final condenser. Seawater, at 71°F, from the power plant cooling water system is used to transfer the latent heat from the condenser to Long Island Sound, along with the waste heat from the power generating facility. Seawater is also used in the product water cooler to lower the temperature of the warm product condensate, in the air ejector system to condense steam, and in the vent system to condense gases.

The remaining 42 percent of the steam (approximately 500,000 lbs/hr) from the last effect is condensed in a barometric leg condenser by raw, ungritted wastewater, raising the temperature of the incoming wastewater from 65°F to 93°F before entering the grit chamber.

The condensate from the final condenser is combined with the condensate from the air ejector system, condenser vents, and that which has accumulated from Effects 2 through 19 to form the product water stream. The combined product stream is at a temperature of 110°F and must have its sensible heat removed in the product water cooler to reduce the product water temperature to 85°F. As discussed previously, seawater is used as the cooling medium.

The cooled product is pumped to an activated carbon absorption system where any residual COD and odor-forming compounds will be removed. Because of the sterilizing environment and phase change in the distillation plant, the final product would not contain any harmful bacteria or viruses.

The quality of the water produced by the integrated facility is anticipated to consistently meet or exceed the water quality criteria established by the U. S. Public Health Service (24) and the World Health Organization (25). The recommended standard of the American Water Works Association (26) will also be met, at all times, during normal plant operation. These various standards are compared in Table 10.

A comparison of the estimated, capital costs for the distillation facility with, and without the provision for heat addition by use of barometric condenser is presented in Table 11.

The distillation plant as described above, would be capable of producing 47.5 MGD of high quality water at an estimated cost of 64¢/1000 gals., which includes the cost of the barometric condenser system for heating the wastewater, and 62.4¢/1000 gals. where no provision is made for wastewater heating.

Of the total distillation plant capital cost, $1,992,000 is attributable to the additional equipment and the increased throughput capacity required in the distillation plant in conjunction with the thermal wastewater treatment process. This incremental capital cost corresponds to an annual fixed charge of approximately $155,000. Of the total annual O&M charges listed in Table 12, $131,000 is attributable to the requirement to transfer heat to the wastewater treatment process. The total annual cost specifically attributable to the thermal enhancement of the wastewater is, therefore, $286,000 which is equivalent to approximately 1.6¢/1000 gals. at a 47.5 MGD production rate. A detailed evaluation of the economics of utilizing steam from the distillation plant to heat the incoming wastewater is presented in Appendix B.

TABLE 10. WATER QUALITY CRITERIA

DRINKING WATER STANDARDS (mg/l except as indicated)

SUBSTANCE	USPHS Conc.-should not be exceeded	AWWA Recommended Goals	WHO Drinking Water Max. Accept.
Turbidity (Units)	5	$<$ 0.1	5
Color (Units)	15	$<$ 3	5
Odor (Threshold Odor No.)	3	no odor	Unobject.
Taste		Nothing object.	Unobject.
Alkyl Benzene Sulfonate (ABS)	0.5	$<$ 0.2	0.5
Aluminum (Al)		$<$ 0.05	
Arsenic (As)	0.01	0.01	
Barium (Ba)		1.0	
Chloride (Cl)	250		200
Calmium (Cd)		0.01	
Chromium (Cr^{+6})		0.05	
Copper (Cu)	1.0	$<$ 0.2	1.0
Carbon Chloroform Extract (CCE)	0.2	$<$ 0.04	0.2
Cyanide (Cn)	0.01	0.01	
Iron (Fe)	0.3	$<$ 0.05	0.3
Lead (Pb)		0.05	
Manganese (Mn)	0.05	$<$ 0.01	0.1
Nitrate (NO_3)	45	45	
Phenols (C_6H_5OH)	0.001	0.001	0.001
Selenium (Se)		0.01	
Silver (Ag)		0.05	
Sulfate ($SO_4^{=}$)	250		200
Total Dissolved Solids (TDS)	500	200	
Zinc (Zn)	5	$<$ 1.0	5.0
Fluoride (F)	1.3	1.3	
Hardness (As $CaCO_3$)		80-100	
Suspended Solids (S.S.)		1.0	
Phosphate ($PO_4^{=}$)		none	
Calcium (Ca)			75
Magnesium (Mg)			50
pH (Value)			7-8.5
Methelene Blue Active Substances (MBAS)	0.5	0.2	
Carbon Alcohol Extract (CAE)		0.15	

TABLE 11. CAPITAL COST DISTILLATION PLANT
(ENR-1690)

		Case I*	Case II**
1.	Decarbonation basin and buffer reservoir	$ 99,000	$ 96,700
2.	Acid and antifoam injection systems	95,400	93,100
3.	Deaeration tower	71,200	69,650
4.	Multiple effect evaporator	56,442,800	53,176,200
5.	Product condenser	1,447,000	3,000,000
6.	Product cooler	1,869,000	1,335,000
7.	Activated carbon adsorption system	1,276,700	1,276,700
8.	Product chlorination system***	117,000	117,000
9.	Intermediate steam reboiler	2,564,000	2,500,000
10.	Intake/outfall structure (amount charged to distillation facility - 70,000 gallons of 740,000 gallons total)	985,000	1,500,000
11.	Ammonia stripper and vent condenser	10,900	10,650
12.	Barometric condenser system	179,000	--
		$65,167,000	$63,175,000

* Case I Distillation plant providing heat to wastewater treatment plant using barometric leg condenser

** Case II No heat added to wastewater treatment plant

***Including provision for chlorinating 50 MGD of secondary treated wastewater in the event of distillation plant outage

TABLE 12. DISTILLATION PLANT WATER PRODUCTION COSTS

		Case I*		Case II**
Distillation Plant Capital Cost (See Table 11)		$65,167,000		$63,175,000
A. Annual Fixed Charges @ 7.823%		5,098,000		4,943,000
B. Annual Operating and Maintenance Charges				
1. Distillation Plant O&M				
Labor	$ 420,000		$ 408,000	
Electric Power	710,000		789,000	
Chemicals	715,000		694,000	
Spare Parts, etc.	918,000		891,000	
Steam	3,082,000		2,991,000	
2. Reboiler O&M	26,000		25,000	
3. Post-treatment O&M	20,000		20,000	
4. Barometric Condensers				
O&M	2,000		--	
Electric Power	56,000		--	
Total Annual O&M		5,949,000		5,818,000
C. Total Annual Charges		$11,047,000		$10,761,000
D. Water Cost - 47.5 MGD Production	$0.64/1000 gal		$0.624/1000 gal	

* Case I Distillation plant providing heat to wastewater treatment plant using barometric leg condenser

**Case II No heat added to wastewater treatment plant

SECTION X

OVERALL PLANT INTEGRATION AND EVALUATION

In accordance with the objectives of this study conceptual designs were developed for an integrated facility incorporating 1000 Mwe nuclear power plant, a 50 MGD wastewater treatment plant, and a 47.5 MGD distillation plant, at a reference site located on the north shore of Long Island in the Town of Riverhead, Suffolk County, New York. As discussed in the preceding sections, the anticipated growth in population, the increased need for electricity and the projected inadequacy of ground water supplies make the concept of integrating electrical, water supply and wastewater treatment facilities attractive in this area. Systems to transport the wastewater from Disposal Districts 11 and 13 (and possibly a portion of the waste from Disposal District 12) to the integrated facility could be planned and installed, well within the time frame required for its activation. The integrated facility concept, in accomplishing the objectives of water reuse and resource conservation satisfies the requirement for tertiary wastewater treatment and meets all present and proposed federal, state, and local standards for protection of the environment. The distillation process, incorporating sterilizing temperatures and phase change, assures the high quality of the product water. Postdistillation treatment of the product water, including activated carbon absorption, mineralization, and chlorination, can be provided as required.

In addition to sharing a site and administrative facilities, the three functional components of the complex are integrated through interties at a number of process points. The nuclear steam supply system, in addition to serving the power generation facility, provides process steam to the distillation plant. The nuclear steam supply and power generating facility will provide all the electrical power for the pumps and other motor driven equipment as well as site lighting, heating, ventilating, and air conditioning, while producing approximately 1000 Mew for off-site transmission.

The distillation plant receives its thermal energy from the nuclear steam supply system, and in turn provides heat for enhancement of the wastewater treatment process by utilizing a barometric leg condenser through which approximately 42 percent of the product steam of the final stage is introduced to the wastewater. (This steam would normally be condensed by the plant cooling water and the heat disposed of in the receiving body of water.) This addition of energy is sufficient to raise the temperature of the wastewater stream, as received, from a nominal 65°F (18°C) to the initial processing temperature of 93°F (34°C).

The distillation plant acts as a tertiary stage of wastewater treatment which eliminates or replaces treatment steps that would be required for discharge to the environment. Ammonia removal is accomplished through venting of the initial stages of the distillation train. Phosphates, nitrates, and refractory organics passing through the secondary treatment stage of the waste treatment plant will also be removed in the distillation process. Chlorination facilities provided for the product water would serve the wastewater treatment plant if the distillation plant were temporarily shut down, eliminating the need for a separate chlorination station. The seawater circulation system, with its elaborate intake and outfall structures required for the turbogenerator condenser, will also provide cooling water for the distillation plant condenser and product water cooler, afford a convenient means of disposing of the sterile concentrates from the distillation plant, and eliminate the need for an outfall system to serve the waste treatment plant during periods of distillation plant outages.

The integrated facility substantially reduces the amount of water removed from and discharged to Long Island Sound as compared with separate facilities. A 50 MGD treatment plant would discharge approximately 50 MGD of treated and chlorinated wastewater to the sound; whereas the integrated facility will discharge no wastewater except during times of distillation plant shutdown. A 50 MGD unintegrated desalination plant would remove approximately 259 MGD of 33,000 ppm (of chloride) feed and cooling water and discharge approximately 50 MGD of 60,000-70,000 ppm (of chloride) brine and 159 MGD of cooling water heated to about 20 degrees above the intake temperatures. The proposed integrated distillation plant will discharge only 2.4 MGD of 15,000 ppm sterile evaporator concentrate and approximately 100 MGD of cooling water.

ALTERNATIVE METHODS OF WASTEWATER TREATMENT

A detailed literature search, conducted to evaluate the effect of elevated temperature on various wastewater treatment processes and operations, is described in Appendix A. It was concluded that existing work is fragmentary in that only single unit operations have been considered and no systematic evaluation has been made of heat input to the overall process.

Nonetheless, the limited information was sufficient to conclude that the operation of conventional treatment units at elevated temperatures can be expected to improve removal efficiencies. Thus, under conditions of thermal enhancement, as contrasted to operation at ambient temperatures, a given throughput may be handled in a plant of smaller physical size, while maintaining the same efficiency and effluent quality.

The processes examined generally fall into the categories of primary, secondary, or tertiary treatment, all of which include sludge treatment and disposal. Unit operations for primary treatment include bar screening, comminution, pumping, grit removal, sedimentation and chlorination.

As an alternative to sedimentation, flotation was considered but not incorporated into the plant design because flotation is generally less effective than sedimentation for removal of solids, although elevated temperatures and steam injection techniques could improve the flotation process.

For secondary treatment, the alternatives considered in lieu of the activated sludge process were the use of trickling filtration and stabilization basins.

For the integrated plant, consideration was given to utilizing a stabilization basin as a cooling pond for the nuclear power system. With this concept, the wastewater temperature would be elevated either by passage through the turbogenerator condenser, or by a separate exchanger through which the condenser coolant would be passed. This arrangement, in which the aeration basin could be eliminated, was rejected because adequate information is unavailable relative to heat transfer performance under potential fouling conditions associated with wastewater processing, as well as the extensive land area which would be necessary for the cooling pond stabilization basin.

Trickling filtration was rejected because effluents from that process have been found to be less desirable than those from activated sludge effluents for feed to distillation units. Additionally, use of trickling filters would, by increasing the rate of heat loss to the atmosphere, require more energy to be transferred to maintain elevated temperatures in the secondary clarifiers. While this requirement is consistent with the objective of dissipating waste heat, the state-of-the-art of heat transfer technology militated against this approach.

In view of the foregoing considerations, therefore, the activated sludge system incorporating primary sedimentation, anaerobic digestion, and vacuum filtration was selected for this project because the activated sludge process is the only treatment system with which there is sufficient experience to be incorporated into a plant of this size.

In order to produce a quality effluent, a tertiary treatment process is employed. The choice among various alternatives for achieving tertiary treatment depends on the specific effluent standards to be met. In the case at hand, distillation was selected as the principal tertiary treatment process in order to assure a high quality product. A conventional activated sludge plant operated with 2,000 to 5,000 milligrams per liter of suspended solids in the aeration tanks and 6 hours detention time will produce an effluent that is suitable for distillation. Carbon absorption was added as a polishing step and a means of insuring the removal of certain volatile contaminants, such as phenols, alcohol and aldehydes which might be present in the wastewater and which could be carried over to produce objectionable tastes and odors in the product water.

The activated sludge treatment process will produce an effluent which will be low in BOD but high in nitrogen, which subsequently will be removed by venting the initial stages of the distillation plant. The vented vapors are processed through a condenser and ammonia stripping unit; whereas, in a separate facility nitrogen removal would be effected by a more costly nitrification-denitrification activated sludge process.

UTILIZATION OF HEAT IN WASTEWATER TREATMENT

The design criteria for the processing units which are incorporated in this plant are developed in Appendix A. The settling velocities, the rate of biological activity, the power required to dissolve oxygen in the wastes and the detention time required for effective disinfection with chlorine are all dependent upon temperature. Once the solids have been separated from the wet stream, the only unit operation affected by temperature is gravity thickening of combined waste, consisting of activated and primary sludges. Digestion will take place in the mesophilic range as would the case in a conventional plant.

Investigators have reported that the optimum temperature for mesophilic biological growth lies in the range of 30°C to 37°C, while a thermophilic growth lies in the 50°C to 54°C range. 30°C and 52°C are generally considered to be the optimum temperatures for mesophilic and thermophilic aerobic biological treatment, while 36°C and 54°C are considered optimum for anaerobic digestion processes. Thermophilic processes were ruled out because insufficient experience exists with the process in large scale operation, and the process is extremely sensitive to changes in temperature.

The choice of mesophilic aerobic process at the temperature of 30°C for the integrated plant was dictated by the apparent leveling off of improvement in efficiency of the activated sludge process beyond that temperature, and by the difficulty in achieving higher temperatures with the available heat sources.

Illustrative of the overall system improvement to be gained, it is projected that increasing the temperature of the wastewater to 30°C (in the aeration tanks) and keeping the unit sizes constant results in an increase in overall treatment efficiency from 90.5 percent to 93.0 percent. Conversely, maintaining the removal efficiencies and effluent quality equal to that achieved by an ambient temperature plant permits significant reduction in unit sizes. For example, the primary settling tank surface area was reduced from 53,600 ft^2 to 42,000 ft^2. Similar results were achieved for the secondary clarifier. The reduction in aeration basin volume was reduced from 2,073,600 ft^3 to 1,695,000 ft^3. A complete comparison of the selected unit operations for the conventional and the integrated, heated 50 MGD treatment plants is shown in Table 13. It should be noted that sludge handling processes such as thermal conditioning and combustion or wet oxidation, which are performed subsequent to gravity thickening, would not be affected by the initial heating of the wastewater. Basically, only the wet stream processes of primary settling, aeration, final settling, gravity thickening and chlorination are affected by temperature.

WASTEWATER TREATMENT HEAT BALANCE ANALYSIS

Heat balances were made on the proposed integrated, heat treatment plant to determine the quantity of the heat lost by each unit in different seasons, and the quantity of heat necessary to achieve the design temperature. The efficiencies of the primary and secondary clarifiers increase with temperature in the range of 10°C to 60°C, while the activated sludge unit reaches constant efficiency at a temperature of 86°F (30°C); hence, a heat balance helps to insure that the plant operates at or near optimal efficiency under a broad range of external conditions.

The factors affecting heat losses include temperature differentials between the wastewater and the atmosphere and ground, surface wind velocity, solar radiation, relative humidity, equipment surface area and type of flow considered, either plug or mixed. These factors were then analyzed for their applicability to the proposed integrated plant and the geographical area. In order to determine the average quantity of heat required to maintain design efficiency, the applicable factors were evaluated under conditions existing during the four seasons of the year. The most extreme conditions occur in the winter months when the temperature differentials between wastewater and the atmosphere and ground are at maximum values. Wind velocities are at a maximum value, while solar radiation is at a minimum value. A statistical approach was used in this analysis, incorporating mean values of winter temperature, wind velocity, relative

TABLE 13

COMPARISON OF UNIT SIZES FOR CONVENTIONAL AND INTEGRATED, HEATED 50 MGD TREATMENT PLANTS

UNIT	UNIT SIZES CONVENTIONAL T = 20°C	INTEGRATED HEATED T = 30°C
Raw Waste Pumping Station	75 MGD	75 MGD
Grit Chambers	L = 61.2 ft	L = 51 ft
Primary Settling Tanks	53,600 ft^2 S.A.	42,000 ft^2 S.A.
Primary Sludge Pump Station	214 gpm	222 gpm
Aeration Basins	2,073,600 ft^3	1,695,000 ft^3
Aeration Equipment	1920 HP	1920 HP
Final Clarifier	66,240 ft^2 S.A.	51,800 ft^2 S.A.
Return Sludge System	50 MGD	52 MGD
Gravity Thickener	11,538 ft^2	8860 ft^2
Anerobic Sludge Digesters	828,990 ft^3	828,990 ft^3
Vacuum Filter Facil-ites	2580 ft^2	2214 ft^2
Chlorine Contact Tank	116,148 ft^3	___*
Chlorine Feed System	4395 #/Day	___*

*-Not necessary in integrated plant (heated or unheated).

humidity and solar radiation. However, on any particular week or day, the mean values may be exceeded and heat losses will rise above the value necessary to maintain maximum efficiency. A factor of safety may be applied if the quantity of heat added is to be sufficient for a critical month, week or day. The risk of not supplying enough heat for a particular period may be weighed against the additional costs and the reduced quality of effluent water. Table 14 summarizes the results obtained from calculations utilizing a computer program developed specifically for this purpose. Average heat input was calculated using values obtained for the four seasons and thus is not representative of the extreme case. The basic values selected for the climatological factors were 15 mph for wind velocity, 10°F for air temperature and 70 percent for relative humidity.

APPROACHES FOR HEAT ADDITION

The method used for the addition of heat to the wastewater treatment process depends in part on the source used to provide the heat. Seven basic sources of heat were identified and considered in this study, including; prime or high temperature steam produced specifically for heating purposes, low pressure or process steam, turbogenerator exhaust steam, power plant and distillation plant cooling water streams, exhaust steam from the final stage of the distillation plant, extracted steam from intermediate stages of the distillation plant, and the distillation plant product water. The alternatives of high temperature and process steam were rejected on the basis of economics, insofar as the cost of high temperature prime steam or even process steam is much greater on a unit heat value basis than low temperature energy, such as spent steam and distillation plant exhaust steam. Furthermore, based on a review of existing literature, relative to the effect of heat on wastewater treatment, no significant advantage was found in raising the wastewater treatment temperature above 93°F.

Heating of the incoming wastewater by countercurrent cooling of warm product water or by condensation of exhaust steam from the turbogenerator appears to be the most attractive approach. The energy from these sources is waste heat normally rejected to the cooling water. However, conflicting

TABLE 14

HEAT BALANCE ANALYSIS

Climatological Basis	---------Yearly Averages---------		
Treatment Plant Temperature	86°F (30°C)	104°F (40°C)	122°F (50°C)
Temperature (°F)			
Raw Wastewater	65	65	65
Grit Chamber			
Influent	93	112	131
Effluent	92	111	130
Primary Settler			
Influent	92	111	130
Effluent	91	110	129
Aeration Tank			
Influent	91	110	129
Effluent	86	104	122
Secondary Settler			
Influent	86	104	122
Effluent	85	103	121
Heat Input (10^8 BTU/hr)			
Heat Exchanger Capacity at Average Flow	4.85	8.2	11.5
Heat Exchanger Capacity at Peak Flow	7.3	11.3	17.2
Condensate Flow MGD at Peak Flow	2.0	3.4	5.1

and insufficient information relative to heat exchange technology precluded the adoption of these energy sources for the design of the integrated facility. In the case of surface heat exchangers, the limited data presently available on wastewater fouling factors and associated heat transfer coefficients, as well as uncertainties regarding the materials of construction and the costs of fabrication, suggest that further information be developed as a prerequisite for considering this approach.

Another alternative considered was the use of the power plant and distillation plant cooling water streams. This approach was rejected because the temperatures of these coolant streams are constrained, by thermal standards, to values that will not produce the 93^{o}F necessary in the wastewater.

The exhaust steam from the distillation plant or extracted steam from the latter stages of distillation constitute the remaining alternatives. Heat from these stages may be transferred to the wastewater stream utilizing either surface heat exchangers of barometric leg condensers. Again, in the case of surface heat exchangers, performance and cost uncertainties must be resolved. In the case of the barometric leg condenser, the direct cost of heat transfer is very low, but high purity product is lost and recycled into the wastewater system. Maintaining the desired production from the distillation plant under these conditions requires an increase in the design capacity and operating throughput of all equipment between the point of heat addition in the wastewater process and the final stage condenser of the distillation train.

A detailed analysis of the engineering and economic factors associated with addition of heat to the wastewater treatment plant was performed and is presented in Apprndix B. This analysis indicated that, based on available information relative to heat transfer surface fouling in wastewater service and expected heat transfer coefficients, utilization of barometric leg condensers to inject a portion of the steam from the final distillation stage into the incoming raw waste would result in the lowest overall cost penalty for the system. For this reason this approach was incorporated in the basic flowsheet for the integrated facility, and resulted in a reduction in the capital cost of the wastewater treatment facilities which was, however, less than the increase in capital cost calculated for the distillation plant.

It should be noted, however, that information provided to ASDA by various manufacturers (27, 28, 29) indicates that higher heat transfer coefficients and lower installed equipment costs than those considered in this study may be obtainable. For a maximum wastewater treatment temperature of 93°F, surface heat transfer equipment would be competitive with barometric leg condensers, if the total installed cost of the former using low grade steam was no more than $2 million. In the case of a product water/wastewater exchanger, the total installed cost could be as much as $3.3 million and ill be competitive. The reduction could be achieved either by a decrease in equipment fabrication or installation costs or by an improvement in the overall heat transfer coefficient resulting in a reduction in the total required heat transfer area.

ALTERNATIVE APPROACHES OF INTEGRATION

A number of alternative approaches of integrating the facilities were considered during the course of this study. One approach which was considered to be highly promising was based on the utilization of partially treated wastewater as the make-up to a closed evaporative cooling system for the power generation and distillation plant facilities. Since the blowdown from this system would provide the feed to the distillation plant, routine discharges to the aquatic environment would be completely eliminated with the exception of the evaporator bottoms which could be further concentrated, dried and disposed of as landfill or by incineration. A spray pond utilizing floating power spray modules was considered to be a feasible approach and a logical choice for evaporative cooling at this site. In addition to discharging all heat from the turbogenerator and distillation plant condensers to the atmosphere rather than to an aquatic heat sink, the spray module approach provides aeration of the partially treated waste and, thus, serves as an additional stage of wastewater treatment. The disadvantage of this approach, and the primary reason it was ultimately rejected, was that the cooling requirements for the facility would result in the evaporative loss of approximately 18 MGD of water which could otherwise be largely returned to the water supply of the region.

Furthermore, Long Island Sound provides an adequate cooling water supply at this site and with proper outfall design could be utilized as a heat sink without adverse environmental impact under existing heat load conditions and thermal discharge criteria. For sites with more restrictive environmental constraints on thermal discharge or lacking in water supply for cooling purposes, the use of partially treated waste for condenser cooling appears to offer significant economic and environmental advantage.

Alternatives to distillation were considered and rejected in the planning stage of this project. Although other desalting techniques, particularly reverse osmosis and electrodialysis, appear to have promise for the recovery of brackish waters or wastewater, these processes are less amenable to integration with a large scale power generation facility and lack the fail-safe characteristics of distillation considered necessary when converting wastewater to a high quality product.

Multistage flash (MSF) and combined multistage flash and vertical tube evaporative (MSF-VTE) systems were investigated in addition to the VTE approach which was finally selected on the basis of recent developments and improvements in the economics of this system (30). The VTE also has greater flexibility than MSF plants both in mode of operation and maintainability. Consideration was also given to operation of the distillation plant solely, or largely, in off power peak periods in order to obtain greater electric power outputs during peak demand periods and permit reductions in cost of steam used for distillation plant purposes. However, the relatively high operating efficiency of the VTE system and operational control considerations led to the conclusion that the plant and the power generating facility should both be operated in a base loaded mode. This approach also minimizes the need for buffer reservoirs between segments of the water treatment and recovery processes.

BENEFITS OF INTEGRATION

Integration of power generation, wastewater treatment, and water supply facilities results in both tangible and intangible benefits. Reuse of water results in conservation of fresh water supplies that would otherwise be lost if the wastewater were treated and discharged to Long Island Sound. The integrated system results in the reduction of the buildup of phosphates, organics, and nitrates observed in areas served by septic systems. Buildup of nitrates occurs in groundwater systems where a substantial portion of the resupply is from treated wastewater and nitrates are considered to cause eutrophication when discharged to certain surface waters. The current advanced water treatment methods for the removal of nitrogen includes ammonia stripping with air, nitrification-denitrification, and breakpoint chlorination. All of these processes have limitations. The ammonia stripping process is severely limited by low temperatures and would be economically prohibitive in a conventional 50 MGD treatment plant. Because heat is supplied in the distillation process, ammonia stripping can be accomplished on a year-round basis. Furthermore, the ammonia need not be released to the atmosphere but can be concentrated and used or converted to nitrogen and harmlessly released to the atmosphere.

Reuse not only creates a water resource, which could be vital to the maintenance of the water balance of the region in the future, but also addresses the recently established criteria for "zero discharge" by eliminating many of the pollutants that would be discharged from separate facilities. The high quality of the water which can be produced by this system would augment water supplies in the area, either through distribution, recharge to groundwater aquifers, or supply for industrial purposes.

Utilization of auxiliary and ancillary facilities, jointly, will result in operational and economic savings. Savings estimated at $3.9 million will result from elimination of separate administrative, chlorination, and outfall facilities for the wastewater treatment plant. Other benefits include reduction in fencing, access roads, number of maintenance and security personnel, and other similar items common to the three plants.

COMPARISON OF INTEGRATED AND CONVENTIONAL SEPARATE FACILITIES

In the absence of the integrated facility the electrical, wastewater treatment, and potable water demands of the region would have to be met by separate conventional power stations, waste treatment plants, and water supply facilities. Additional electrical energy production, considering the projected growth in population and electrical demand in the region, would likely be provided by large nuclear power generating stations similar to that proposed for the integrated facility without the capability for dual purpose operation. A recent estimate (15) of the cost of such a facility (1972 dollars, ENR index 1690) is approximately $284/Kw or $284 million for 1000 Mwe unit. The calculated bus bar electrical cost for such a unit if built and operated with once-through seawater cooling, is approximately 9.1 mills/Kw-hr.

The total projected capital cost for the nuclear steam supply and power generating portion of the integrated facility is $301.5 million with the increase in cost attributable to the larger steam supply system required for the dual-purpose operation. However, since in the integrated facility steam will be "sold" at cost to the distillation plant, the bus bar electrical energy costs remains unchanged at 9.1 mills/Kw-hr.

The capital and operating costs for a separate 50 MGD conventional activated sludge wastewater treatment plant, including outfall structure, chlorination facilities, and administrative facilities are presented in Tables 15 and 16. The $25.7 million capital cost of secondary wastewater treatment at such a facility are equivalent to approximately 16.2¢/1000 gallons assuming a fixed charge rate 7.842%.

The capital cost allocated to the wastewater treatment portion of the integrated facility totals $20.3 million and is detailed in Table 17. The operating costs for such a plant are estimated to be $861,000 annually and are detailed in Table 18. The wastewater treatment costs, including fixed charges and operating costs are equivalent to 13.5¢/1000 gallons.

The estimates in Tables 15 and 17 for wastewater pumping reflect several pumping stations located in the sewage

TABLE 15

50 MGD WASTEWATER TREATMENT PLANT COST ESTIMATE
CONVENTIONAL SYSTEM (NONINTEGRATED, UNHEATED)
(ENR 1690)

Raw Water Pump Station	$ 2,415,000
Grit Chambers	685,000
Primary Clarifiers	1,005,000
Primary Sludge Pumping Station	120,000
Aeration Basins	2,750,000
Aeration Equipment	1,280,000
Final Clarifiers	1,300,000
Return Sludge System	380,000
Gravity Sludge Thickeners	400,000
Anaerobic Sludge Digesters	2,415,000
Vacuum Filter Facilities	1,790,000
Yard Piping	2,060,000
Chlorination Facilities	405,000
Administration Building	160,000
Maintenance Facilities	160,000
Outfall	2,415,000
Subtotal	19,740,000
Engineering and Contingencies @ 30%	5,922,000
TOTAL	$25,662,000

TABLE 16

50 MGD WASTEWATER TREATMENT PLANT OPERATING COSTS
CONVENTIONAL SYSTEM (NONINTEGRATED, UNHEATED)

A.	Annual Fixed Charge @ 7.823% of $25,662,000 =		$2,008,000
B.	Annual Operating and Maintenance Charges		
	1. Labor	$ 320,000	
	2. Power	172,000	
	3. Chemicals	62,000	
	4. SUBTOTAL	554,000	
	5. Miscellaneous @ 10% (1 & 2 & 3)	56,000	
	6. SUBTOTAL	610,000	
	7. Contingencies @ 10%	60,000	
	8. SUBTOTAL	670,000	
	9. Maintenance Budget	270,000	
	TOTAL Annual O & M		940,000
C.	TOTAL Annual Charges		2,948,000
D.	Water Costs - 50 MGD Treatment		16.2¢/1000

TABLE 17

50 MGD WASTEWATER TREATMENT PLANT COST ESTIMATE
INTEGRATED SYSTEM, PLANT HEATED BY BAROMETRIC CONDENSER*
ENR 1690

Raw Waste Pumping Stations	$ 2,415,000
Barometric Condenser	Included in distillation plant cost-see Chapter IX
Grit Chambers	580,000
Primary Settling Tanks	810,000
Primary Sludge Pump Station	130,000
Aeration Basins	2,240,000
Aeration Equipment	1,280,000
Final Clarifiers	1,055,000
Return Sludge System	395,000
Gravity Thickeners	340,000
Anaerobic Sludge Digesters	2,415,000
Vacuum Filter Facilities	1,790,000
Maintenance Facilities	160,000
Yard Piping	2,000,000
Subtotal	$15,610,000
Engineering and Contingencies @ 30%	4,685,000
Total	$20,295,000

*-Inlet to grit chamber, 93°F; primary settling tanks, 92°F; aeration basins, 86°F (30°C); plant outlet, 85°F.

TABLE 18

50 MGD WASTEWATER TREATMENT PLANT ANNUAL OPERATING COSTS INTEGRATED SYSTEM, PLANT HEATED BY BAROMETRIC CONDENSER

A.	Annual Fixed Charge @7.823% of 20,295,000		1,588,000
B.	Annual Operating and Maintenance Charges		
	1. Labor	300,000	
	2. Power	172,000	
	3. Chemical	0,000	
	4. Sub Total	472,000	
	5. Misc.@10% (1&2&3)	47,000	
	6. Sub Total	519,000	
	7. Contingencies @10%	52,000	
	8. Sub Total	571,000	
	9. Maintenance	290,000	
	Total Annual O & M		861,000
C.	Total Annual Charges		2,449,000
D.	Water Costs - 50 MGD Treatment		13.5¢/1000 gal.

districts to convey the wastewater via forced main to the facility site and, thereby, preclude the need for pumping facilities on the site.

The primary sludge and return sludge pumping stations and maintenance building sizes and costs were adjusted to reflect the 3 percent increase in flow due to the steam condensate from the barometric condenser used to supply heat to the wastewater.

To operate the aeration basins at 86°F (30°C) the temperature of the water entering the grit chamber is 93°F and in the primary clarifiers is approximately 92°F. Reductions in grit chambers and primary clarifier costs reflect operation at these temperature levels. The required horsepower to force oxygen into the aeration tanks is reasonably constant over the temperature range considered. Although the ability to transfer oxygen increases with temperature, oxygen saturation concentration decreases simultaneously and the two effects cancel.

Cost data for an integrated, unheated, 50 MGD plant are given in Table 19. In this case, chlorination facilities, an outfall structure and an administration building are all eliminated. Disinfection, by chlorination is not required since the process following final waste treatment is distillation. The outfall is not required because the entire treated effluent flow is passed through the distillation plant, or in the event of a distillation plant shut down, through the outfall of the power plant cooling water system. The administrative facilities of the power plant-distillation plant complex are considered to be adequate to service the needs of the wastewater treatment plant.

Comparison of the conventional nonintegrated, unheated plant and the integrated, unheated plant cost estimates indicates that savings in treatment plant construction costs on the order of 15 percent are possible when the wastewater treatment facility is incorporated to the power plant-distillation plant complex. This saving is due to integration alone and the effect of heat addition on the cost saving is not included.

By comparing the capital costs for the integrated unheated

TABLE 19

50 MGD WASTEWATER TREATMENT PLANT COST ESTIMATE
INTEGRATED SYSTEM, PLANT NOT HEATED
ENR 1690

Total Cost, Conventional System		$25,662,000
Cost of items to be eliminated:		
• Chlorination Facilities	$ 405,000	
• Administration building	160,000	
• Outfall	2,415,000	
Subtotal	$2,980,000	
• Engineering & contingencies	890,000	
Total Savings	$3,870,000	
Total Cost, unheated, integrated system		21,792,000
Net Benefit of integration		3,870,000
Percentage Cost Reduction due to integration (Basis-Total Cost, Conventional System)		15%

and heated plant, it is readily apparent that a reduction of $1,497,000 is possible. This figure represents an additional 5 percent reduction in the construction cost of the wastewater treatment facility, over and above the 15 percent realized by integration. However, if the cost associated with the addition of heat to the wastewater, which is primarily reflected in increased capital and operating costs for the distillation plant, is charged to the wastewater treatment plant, the cost is increased by approximately 1.6¢/1000 gallons, for a total cost of 15.1¢/1000 gallons.

Table 20 shows the effect upon capital costs by adding heat to the wastewater treatment facility by means of a shell and tube heat exchanger. The major cost difference, as compared to the barometric condenser case, is associated with

TABLE 20

50 MGD WASTEWATER TREATMENT PLANT COST ESTIMATE
INTEGRATED SYSTEM,
PLANT HEATED BY SHELL & TUBE HEAT EXCHANGER
ENR 1690

Total Treatement Plant Cost, Barometric Condenser System	
From Table 16	$20,295,000
Cost additions, using shell and tube heat exchange:	
1. Heat exchange downstream of grit chamber use conventional grit chamber cost	
Cost addition: 685,000-580,000 = $105,000	
Cost reductions, reflecting 3% flow reduction due to removal of barometric condenser:	
1. Primary clarifiers: 0.03 x 810,000 =	25,000
2. Aeration basins: 0.03 x 2,240,000 =	67,000
3. Secondary clarifiers: 0.03 x 1,055,000 =	32,000
Total Cost Reductions	124,000
Net cost reduction, including 30% contingency:	$ 25,000
Total Treatment Plant Cost, Shell and Tube System:	$20,270,000

the grit chamber. Because of potential erosion of the heat exchanger, the sewage must be degritted prior to being heated. Thus, the cost for the grit chamber in the conventional unheated plant is identical to the value shown in Table 15, and increases the cost by $105,000. Against this increase in cost, a 3 percent reduction in cost of the wet stream plant size is possible with the elimination of the barometric leg condenser. The saving in cost resulting from this size reduction is $124,000, which after deducting the $105,000 increase in the grit chamber cost and applying

TABLE 21

CAPITAL COST - 50 MGD INTEGRATED FACILITY HEAT ADDED BY HEAT EXCHANGER

Temperature	$20^{\circ}C$	$30^{\circ}C$	$40^{\circ}C$	$50^{\circ}C$
Raw Waste Pump Station	$2,415,000	$2,415,000	$2,415,000	$2,415,000
Grit Chamber	685,000	685,000	685,000	685,000
Primary Clarifier	1,005,000	785,000	710,000	620,000
Primary Sludge Pump Station	120,000	130,000	143,000	150,000
Aeration Basin	2,750,000	2,173,000	2,244,000	2,276,000
Mechanical Aeration	1,280,000	1,280,000	1,280,000	1,280,000
Secondary Clarifier	1,300,000	1,023,000	900,000	760,000
Return Sludge & Pump Station	380,000	395,000	408,000	418,000
Thickener	400,000	340,000	275,000	260,000
Anaerobic Digestion	2,415,000	2,415,000	2,415,000	2,415,000
Vacuum Filter	1,790,000	1,790,000	1,790,000	1,790,000
Maintenance Building	160,000	160,000	167,000	177,000
Yard Piping	2,060,000	2,000,000	2,136,000	2,133,000
Subtotal	16,760,000	15,591,000	15,568,000	15,379,000
Eng. & Contingencies (30%)	5,032,000	4,679,000	4,670,000	4,614,000
TOTAL	$21,792,000	$20,270,000	$20,238,000	$19,983,000

the 30 percent contingency factor used in preparing the plant estimates yields an additional saving of $25,000, exclusive of the cost of the heat exchanger.

Table 21 demonstrates the relationship between capital costs and the temperature at which the plant is operated, again with heat transferred by means of shell and tube heat exchangers. The importance of the size of the aeration basin in this analysis is evident, particularly at the 30°C level where the aeration basin cost is at a minimum. Clearly, the cost saving to be achieved for temperatures in excess of 30°C are relatively inconsequential compared to the saving associated with the 30°C operation.

Table 22 summarizes the foregoing discussions of various alternatives for integrating the wastewater plant and shows that a 15 percent reduction in the construction cost of the wastewater treatment facility is possible through integration, with an additional 6 percent reduction in cost available through heat addition.

TABLE 22

COMPARISON OF COSTS OF VARIOUS WASTE TREATMENT PLANT CASES

Plant and System	Capital Cost for 50 MGD Plant (In Million $)	Percentage Reduction Over Conventional Plant
Conventional Separate Plant, Unheated	25.66	--
Integrated System, Unheated	21.79	15
Integrated System, Heat Via Barometric Condensers	20.30	21
Integrated System, Heat via Shell and Tube Exchanger	20.27	21

Comparison of wastewater treatment costs, however, must take into consideration the fact that conventional secondary treatment with discharge to the environment is no longer acceptable. Tertiary waste treatment is required to approach the "zero discharge" criterion. The capital cost of wastewater facilities providing tertiary treatment is estimated to be more than twice the cost of facilities providing secondary treatment only and the total annual operating costs for these facilities are estimated to be approximately three times the cost of present conventional waste treatment systems (31). On this basis a separate waste treatment plant or plants serving the study region and designed to meet the water quality goals incorporated in the Federal Water Pollution Control Act Amendments of 1972 (32) would cost in excess of $50 million and the cost of the wastewater treatment would be on the order of 55¢/1000 gallons. Elimination of the need for tertiary waste treatment provides an additional benefit for the integrated facility approach.

An alternate approach to meeting the water demands of the region is seawater desalting. The projected economics of this process are indicative of an advantage for large scale plants similar in size to that proposed for the integrated facility, i.e., in the range of 40 to 60 MGD. In order to achieve operating economy, these plants will be coupled to a dual-purpose power generating facility, most probably nuclear fueled. It is, of course, possible for a desalting plant to have its own thermal energy source. Table 23 presents the capital cost of such a nonintegrated 50 MGD distillation plant. The present and projected costs of fossil fuels, however, indicate that even the most efficient single purpose fossil plants would incur a cost penalty of at least 25¢/1000 gallons over a dual purpose nuclear facility.

Present projections of the cost of large seawater desalting plants coupled with nuclear power generating facilities such as the 40 MGD plant at Diablo Canyon, California (capital cost of $92 million, and product water cost of 92¢/1000 gallons) or the pair of 20 MGD flash train evaporators at Encina, California (capital cost of $72 million and product water cost of 99¢/1000 gallons) (34) illustrates the potential savings associated with an integrated facility having a distillation plant capital cost of $63.2 million, and product water cost of 62¢/1000 gallons.

TABLE 23

CONVENTIONAL NONINTEGRATED 50 MGD VTE DISTILLATION PLANT

Unit Operation	Cost
Steam & Power Supply-Fossil	$ 17,100,000
Distillation Plant	54,200,000
Intake/Outfall Structure	6,100,000
Product Steam Condenser	3,000,000
Product Water Cooling	1,700,000
Pretreatment	270,000
Activated Carbon	1,270,000
Post Treatment	200,000
	$ 83,840,000

The cost for the integrated distillation plant includes the cost of product post-treatment and the full cost of chlorination and administrative facilities shared with the wastewater treatment plant, but excludes the cost directly applicable to the requirement for thermal enhancement of the wastewater treatment process. The unit process cost of 62¢/1000 gallons includes 17.7¢/1000 gallons ($3.1 million per year) for steam purchased from the power facility.

In summary, the total capital cost of the integrated facility is estimated to be $386 million with the total annualized cost estimated to be $77 million per year.

The facility is designed to be capable of producing 7 billion Kw/hr per year of electricity at a cost of 9.1 mills/Kw-hr and 47.5 MGD of high quality product water at 62¢/1000 gallons, while treating 50 MGD wastewater at a cost of approximately 15.1¢/1000 gallons.

Summaries of the capital and annual cost of the integrated facility and the revenues required to recover these costs are presented in Tables 24 and 25.

TABLE 24

INTEGRATED FACILITY COST SUMMARY

I. CAPITAL INVESTMENT (ENR 1690)	$\$10^6$
A. Nuclear Steam Supply and Power Generation	
Nuclear Steam Supply @ $175/Kwe	$192.5
Turbogenerator @ $109/Kwe	109.0
	$301.5
B. Waste Treatment Plant	
Basic Plant	$ 20.3
Heat Addition	2.0
	$ 22.3
C. Distillation Plant	
Basic Plant	$ 61.9
Product Post-Treatment	1.3
	$ 63.2
TOTAL CAPITAL COSTS	$386.0
II. ANNUAL COSTS	
A. Steam Supply and Power Generation	
Fixed Charges @ 15.0%	$ 45.2
Fuel	16.2
Operation and Maintenance	5.3
Less: Proceeds from sale of steam to distillation plant	-3.1
	$ 63.6
B. Wastewater Treatment	
Fixed Charges @ 7.823%	$ 1.75
Operation and Maintenance	.99*
	$ 2.74*
C. Distillation Plant	
Fixed Charges @ 7.823%	$ 4.94
Operation and Maintenance	5.82**
	$ 10.76**
TOTAL ANNUAL COSTS (excluding interutility purchases)	$ 77.2

*Includes $130,000 distillation plant costs for wastewater heating.

**Includes $3,082,000 for steam "purchased" from the power facility.

TABLE 25

INTEGRATED FACILITY
PRODUCTS AND REVENUES

		$\$10^6$
Electricity		
7×10^9 Kw/-hr/yr	@ 9.1 mills/Kw-hr	$ 63.6
Potable Water		
47.5 MGD	@ 62¢/1000 gallons	$ 10.8
Treated Waste		
50 MGD	@ 15.1¢/1000 gallons	$ 2.8
TOTAL REVENUES		$ 77.2

SECTION XI

PILOT SCALE DEMONSTRATION PROJECTS

As discussed in the preceding sections and the appendices to this report, limited design and performance data presently exist on the heating of wastewater, the effects of heat on wastewater treatment processes, and the use of distillation processes to produce potable water from treated wastewater. In addition, public health considerations would preclude the direct recycle of wastewater, even using the distillation process, until the ability to produce high quality water continuously and reliably is demonstrated. For these reasons, a phased program for the development and demonstration of the required technology is recommended prior to implementing plans for a full-scale integrated facility complex. This program will consist of three phases:

- Phase I - Component Tests and Laboratory Screening Studies
- Phase II - System Performance Tests and Demonstrations
- Phase III - Integrated Pilot Scale Demonstrations

Phase I, Components Tests and Laboratory Screening Studies, will be directed to developing the fundamental design and performance data with respect to components and subsystems needed to validate the assumptions and theoretical predictions made in this study and to provide a basis for the subsequent system tests and pilot scale demonstrations. In Phase II, System Performance Tests and Demonstrations, a distillation plant will be combined with heated wastewater treatment and prototype heat exchangers to simulate an integrated system at the 50,000 gpd scale. Phase III, Integrated Pilot Scale Demonstrations, will consist of operating a heated wastewater treatment plant and a distillation plant at an existing sewage plant site. To be meaningful, the Phase III program should be based on an integrated plant having a capacity of at least 500,000 to 5,000,000 gallons per day.

REQUIREMENTS FOR DESIGN AND OPERATING DATA

The objectives and goals of the recommended development and demonstration program can best be defined in terms of the subsystems that will comprise the integrated facility complex and the currently available and required technology for the design of these subsystems.

Nuclear Power Plant

As discussed in prior sections, the principal change in the nuclear power plant will consist of providing the additional thermal output needed to provide the energy required for heating the wastewater and for the distillation plant. With electrical output in the 1000 Mwe range, and with wastewater and product water flows in the 50 MGD range, the required additional output is only approximately 10 percent, which can be provided by selecting currently available reactor designs and matching them with available power generation equipment. No special development or demonstration projects will be required.

Radioactive materials will be present in the steam systems of boiling water reactors and, to a lesser extent, in pressurized water systems. For the reasons discussed in Section VI, a reboiler is used to provide isolation of this steam from the wastewater and distillation plants. The use of reboilers for other purposes is a common practice and thermal and hydraulic design can be considered to be state-of-the-art.

The other aspect of power plant integration to be considered is the possible use of the condensers for wastewater heating, which is discussed later.

Heated Wastewater Treatment

As discussed in Section VIII and Appendix A, heating of wastewater improves treatment processes by increasing the rates of biological activity and improves sedimentation by decreasing the density and viscosity of the wastewater. The review of existing literature relative to sewage treatment at elevated temperatures indicates that fully conclusive data are not available with respect to the quantitative or qualitative effects of heat addition. This is exemplified by the narrow band of elevated temperature experience and the limited references to such qualitative effects as changes in the oxidation-synthesis relationships which control biological treatment.

The limited quantity of available literature has constrained the work described herein to the use of conservative estimates of the reductions in size of conventional wastewater treatment operations resulting from heat addition. Additionally, literature data have not been sufficient to provide a basis for quantitative process designs for treatment unit operations other than those employed in conventional biological treatment flow sheets.

Most significantly, reports appearing in the third quarter 1972 literature of qualitative changes in biological treatment performance due to heat enrichment, and operating experience with elevated temperature biological treatment of soluble industrial wastes, strongly indicates that greater size reductions are possible than were considered for the reference design, and that continued analytical work be undertaken to define the quantitative and qualitative effects of thermal enrichment in wastewater treatment operations.

Heating of Wastewater

As discussed in Sections I and XII and Appendix B, there are three potential sources of energy for wastewater heating:

Power plant condensers

Distillation plant product water cooling

Injection of steam from low pressure stages of the distillation plant

Of these energy sources, heating of wastewater in the power plant condensers or the distillation plant product water coolers appears to be the most attractive, since cooling water would normally be required to remove the waste heat from these sources.

Using waste heat from either the power plant condensers or the distillation plant product water coolers will require the use of surface type heat exchangers. Data on fouling factors and attainable heat transfer coefficients, the available temperature differences, and the materials that must be used to resist corrosion must be developed in order to evaluate the options involving the use of heat exchangers.

The use of injection steam from the low pressure stage using barometric leg condensers is attractive in that the heat exchange equipment costs are low, fouling problems are avoided, and a system of this type could be built based on present technology. This approach is economically limited in that the size of the distillation plant, the flow through the wastewater treatment plant, and the total energy required are all increased. The cost of using this approach is estimated to be less than that of surface type wastewater heat exchangers, conservatively designed on the basis of present technology and the very limited existing data. Pursuing this approach further is not recommended because advances in technology cannot be expected to improve significantly the economics of this approach.

The principal area in which productive technological opportunities exist which could make heated wastewater treatment more economically attractive is the development of reliable and economic wastewater heat exchangers suitable for use in transferring waste heat from the power plant condensers or the distillation plant product water cooler.

The most potentially productive opportunities lie in:

1. Increasing the heat transfer coefficients through material selection and the use of coatings and additives to inhibit scaling and fouling.

2. Maintaining desired heat transfer rates through periodic mechanical or chemical cleaning of the heat transfer surfaces.

3. Developing heat exchanger designs of high reliability and ease of maintenance. Such designs would explore the advantages of higher flow velocities and flow turbulence in enhancing heat transfer.

Distillation of Treated Wastewater

The distillation plant design presented in Section IX is based on the vertical tube evaporation process developed for the desalination of sea and brackish water. Although development and demonstration work is continuing to improve the performance, efficiency, and economics of this process, the VTE process can be considered state-of-the-art. Further, multistage flash evaporation and the other existing distillation desalting processes could also be used with wastewater. Work is also underway on the use of membrane processes for the treatment of wastewater. Because of fouling, biological growth, and the possibility of undetectable failures, it is felt that membrane processes should not be considered for direct reuse applications.

Even though the equipment and processes proposed for wastewater distillation can be considered to be state-of-the-art, there are a number of uncertainties with respect to distilling secondary treated wastewater. With the activated sludge process, a large portion of the nitrogen in the secondary treated effluent will be in the form of ammonia. Since ammonia has a higher vapor pressure than the water in which it is dissolved, it will be evaporated with the water vapor in the distillation process. From a theoretical standpoint and with proper selection of operating parameters, it should be possible to remove this ammonia in a gaseous form with the noncondensible gases. Once removed, the ammonia would be condensed for removal from the process. This will require a separate system for handling the condensate from the air ejectors and may require special materials of construction. For these reasons, the ammonia removal process and equipment should be demonstrated.

Other materials potentially present are phenols and other volatile organic materials which have boiling and condensing temperatures and pressures similar to water or are volatile and highly water-soluble. Such materials could be carried over with the product water, but at low concentrations could be removed inexpensively by carbon absorption as a post-distillation treatment process.

It will be necessary to have extensive distillation process operating and performance data to show extremely high reliability before direct product water reuse could be considered. Even in cases where the product water is for use in groundwater recharge, industrial processes or as makeup for evaporative power plant cooling systems such as cooling towers, a high degree of process control and product purity will be achieved.

PHASE I COMPONENT TESTS AND LABORATORY SCREENING STUDIES

The Phase I Component Tests would consist primarily of bench scale experiments concentrating on the development of basic data required to design and build prototype equipment. In the case of the distillation system, there is minimal need for component testing and work in this area can concurrently proceed with the Phase II system demonstrations using wastewater from a conventional secondary treatment plant. This approach would allow a longer demonstration period for the distillation plant and the accumulation of reliability and maintainability information.

Laboratory Screening Studies of Wastewater Treatment Unit Operations

Theory indicates that thermal enrichment of wastewater could reduce the size and cost or improve the efficiency of wastewater treatment by at least 20 percent. Very recent operating experience in treating soluble industrial wastes at elevated temperatures indicates much higher percentages may be attainable.

The degree to which wastewater treatment processes can be improved must be demonstrated. The Phase I laboratory scale screening studies of candidate unit operations are designed to achieve the following purposes:

1. Determine actual performance characteristics of activated sludge treatment operations at the elevated temperatures considered in this analysis.

2. Evaluate the qualitative changes in the nature of the activated sludge treatment process at elevated temperatures.

3. Determine the performance characteristics of non-biological treatment operations at elevated temperatures, for example, chemical treatment and chemical sludge disposal.

The unit operations to be screened will include those constituting the 50 MGD plant flow sheet developed in this study and additional unit operations that might, based on new information, be indicated to improve upon this flow sheet. The unit operations to be tested include grit removal, settling, activated sludge, chlorination, thickening, digestion and dewatering.

Two series of screenings tests will be performed. The first series will be batch analyses to determine individual unit operation responses. With the information gained through these tests, a continuous, in line process will be set up to model the proposed demonstration project and to determine the interaction of the various processes.

Laboratory scale results will be analyzed in terms of performance and economic implications to identify a priority ordering of unit operations for pilot plant verification, and to project the process benefits and cost savings available through heat addition to sewage processes. Information relevant to the optimum process temperatures for the pilot plant will be developed and potential operational problems that might hamper treatment at elevated temperature identified.

Wastewater Heating

The Phase I Component Test Program on wastewater heating consists of a series of surface heat exchanger heat transfer experiments using wastes of various concentrations over the temperature and flow range of interest. For these tests, it is recommended that electrically heated tubes be used in conjunction with the portable heat transfer and fouling test equipment of the type developed and used by the Heat Transfer Research Institute. This equipment permits the variables which effect heat transfer and fouling to be investigated. The tubes used in these heat transfer experiments will also be used as corrosion test samples and will be examined metallurgically following the tests. The variables to be evaluated include:

1. Degree of wastewater treatment (primary and secondary) needed prior to heating.

2. Temperatures from 65°F to 110°F and 150°F to 300°F.

3. Flow velocities from 2 to 30 feet per second.

4. Tube materials consisting of carbon steel, stainless steel, admiralty brass and other alloys.

5. Inhibitors for corrosion and fouling.

6. Descaling solutions.

In addition to the experimental program, parallel design and application studies will be conducted on surface heat exchange equipment. These studies will review and analyze various heat exchanger configurations for use in an integrated plant.

PHASE II SYSTEM PERFORMANCE TESTS AND DEMONSTRATIONS

Distillation Plant System Demonstrations

The distillation plant system demonstrations will be conducted using equipment developed for use with seawater or brackish water. Ideally, equipment already in the possession of the the Office of Saline Water, U.S. Department of the Interior would be obtained on loan for this purpose. The distillation plant will require a source of steam and secondary treated wastewater. A 50,000 gpd distillation plant for instance will require about 170 MBtu per day, or approximately 7000 pounds of steam per hour, which is the equivalent of a 200 hp packaged boiler using about 50 gallons of fuel oil per hour.

The distillation plant system demonstration could alternatively be located at an existing power plant near an existing wastewater treatment plant. A pumping station could be used to convey the treated wastewater to the distillation plant and the plant could use extraction or low pressure steam from the power plant. Another alternative would be location of the distillation plant where it can be supplied with secondary treated wastewater and use of a packaged boiler.

The distillation plant will require some ancillary subsystems for use with wastewater, including provisions for ammonia removal and for post-distillation treatment. The ammonia removal system will consist of separate steam air ejectors for the removal of noncondensibles and ammonia vapor from the initial stages, an after-condenser, and a small ammonia stripping tower for removal of the ammonia from the condensate. The post-treatment subsystem may consist of parallel carbon adsorption columns with provision for thermal or chemical recycling if phenols are present and not otherwise removed.

The test installation would be supported by a water analysis laboratory, probably the one utilized in the Phase I work.

The test and demonstration program for the distillation plant would consist of a series of test runs to attain operating and performance data, periodic inspections to determine material compatability, corrosion rates, etc., and special tests to simulate process malfunction.

Specific items to be evaluated would include:

1. Operation at various temperatures.

2. Variable bottoms recycle rates and concentrations.

3. Variable feed/product ratios.

4. Use of acid feed.

5. Use of various methods of pretreatment.

6. Descaling methods.

7. Startup, shutdown, and partial load conditions.

8. Water quality as a function of production.

9. Effect of off-standard conditions on water quality:

   a. System flooding;

   b. Loss of vacuum;

   c. Loss of cooling water; and

   d. Loss of feed.

10. Post-distillation removal effectiveness.

11. Carbon column recycle times.

PHASE III INTEGRATED PILOT SCALE DEMONSTRATIONS

The demonstration will be conducted on a scale adequate to permit extrapolation to full scale. Thus the capacity should be at least 500,000 gpd and, preferably, on the order of 5,000,000 gpd to approach within a factor of 10, the designs considered in this study. In this phase, heat addition and related facilities and equipment will be designed and installed using the design and operating data developed. The demonstration will involve addition of heating capability to a soon to be constructed municipal wastewater treatment plant. Heat will be provided from a complex of power facilities located on immediately adjacent land.

The Phase III demonstration will provide an actual operational evaluation of heated wastewater processing. Like a full scale facility, only a limited number of tests will be conducted to validate the design and performance predictions. The major emphasis in the Phase III demonstration will be the collection of performance, reliability, and maintenance data and treated water quality information.

SECTION XII

ACKNOWLEDGEMENT

The support of the project by the Water Quality Office of the Environmental Protection Agency and the guidance and assistance provided by Mr. Patrick M. Tobin and Mr. John R. Trax, the Agency's Grant Project Officers, is gratefully acknowledged.

SECTION XIII

REFERENCES

1. U.S. Bureau of the Census, "Statistical Abstracts of the United States: 1970," (91 Edition) Washington, D.C.

2. Bowe, Walsh & Associates, "Comprehensive Sewerage Studies, Five Eastern Towns, Suffolk County, New York," WPC-CS-158, (Preliminary Draft).

3. New York State Office of Planning Services, Unpublished preliminary population projections, July 1972.

4. Holzmacher, McLendon & Murrell, "Comprehensive Public Water Supply Study," Suffolk County, New York, CPWS-24, (3 Volumes) 1968-1970.

5. Division of Water Resources State Conservation Dept. for State Office of Planning Coordination, "Long Island Water Resources," January 1970.

6. County of Nassau, State of New York, Greeley and Hansen Engineers, "Comprehensive Public Water Supply" (CPWS-60) June 1971.

7. Wiggins, Thomas H., "Report on Comprehensive Plan for the Development and Distribution of the Available Water Supply of Suffolk County, Long Island, New York," Suffolk County Water Authority, January 1957.

8. McGraw-Hill, Electrical World Magazine, September 15, 1970.

9. Battelle Memorial Institute, "A Review and Comparison of Selected United States Energy Forecasts," December 1969.

10. Long Island Lighting Company, "Population Survey 1969- Current Population Estimates for Nassau and Suffolk Counties"

11. Statistical Department, Long Island Lighting Company, "Population Survey 1964."

12. United States Code of Federal Regulations, "Criteria for the Siting of Nuclear Power Facilities," Title 10, Part 100.

13. Nassau-Suffolk Regional Planning Board, "Existing Land Use," 1968.

14. "Preliminary Safety Analysis Report for Surfside Multipurpose Nuclear Reactor Facility," (Unpublished).

15. Vann, Harold E., "Cost Trends for Nuclear Power Plants," Nuclear News/Geneva III, pp. 31-34, October 15, 1971.

16. Hauser, L. G., and R. F. Potter, "The Effect of Escalation on Future Electric Utility Fuel Costs," Nuclear Fuel Division, Westinghouse Electric Corporation, 1971.

17. Leung, Paul, "Cost Separation of Steam and Electricity for a Dual Purpose Station," Bechtel Corporation, San Francisco, California, April 1972.

18. Hittman Associates, Inc., Study for EPA Pacific Northwest Laboratories, "Nomographs for Thermal Pollution Control Systems," (to be Published).

19. Long Island Lighting Company, "Environmental Impact Evaluation for the Shoreham Nuclear Power Station," December 1971.

20. State of Maryland Dept. of Health & Mental Hygiene Regulations for the Control of Air Pollution.

21. Dynatech R/D Company, Report for Water Quality Office EPA, "A Survey of Alternate Methods for Cooling Condenser Discharge Water-Large Scale Heat Rejection Equipment," July 1969.

22. Strobel, J. J., "Overall Economic Considerations of Desalination Operations," OSW, St. Croix Conference, December 1971.

23. "Diablo Canyon Prototype Desalting Plant Design Study," Kaiser Engineers Report No. 70-60-RE, Dept. of Water Resources and OSW, California, January 1972.

24. U.S. Dept. of Health, Education, and Welfare, Public Service Commission, "Public Health Service Drinking Water Standards," 1962.

25. World Health Organization, Geneva, Switzerland, "International Standard for Drinking Water," 1963.

26. American Water Works Association, Inc., Water Quality and Treatment, 1971.

27. Persona l correspondence with Roma Bhattacharya, Product Specialist Heat Transfer Dept., Bell & Gossett and Wallace Eannace Asso ates, Inc.,Bell & Gossetts' New York Representatives.

28. Personal correspondence with William R. Ryan, Sales Manager Water Products & Environmental Systems, Westinghouse Electric Corp. Power System, Heat Transfer Division.

29. Personal correspondence with Anthony J. Giuriceo, Sales Engineer, Mesco Tectonics, Inc.

30. Hittman Associates, Inc., HIT-524 (draft to be Published) "Vertical Tube Distillation Desalting State-of-the Art 1971."

31. Metzler, D.F. and F.O. Bogedain, "The Cost of Water Quality Goals," National Symposium on Cost of Water Pollution Control, April 1972.

32. Federal Water Pollution Control Act Amendments, 1972, Public Law 92-500, Title Three.

33. Office of Saline Water, "Draft Environmental Impact Statement for Diablo Canyon Prototype Desalting Plant," April 1972.

34. Water Desalination Report, Richard Arlen Smith, Publisher-Editor, P.O. Box 35-K, Tracey's Landing, Md. 20869. Vol. III, No. 36.

SECTION XIV

LIST OF PATENTS AND PUBLICATIONS

No inventions, patents or publications have resulted from the performance under or in conjunction with this contract.

SECTION XV

GLOSSARY OF TERMS, ABBREVIATIONS

TERMS

Coagulation	The process of agglomeration of small particles into larger particles through agitation with or without the aid of chemicals.
BOD	Biochemical Oxygen Demand.
COD	Chemical Oxygen Demand.
Floc	A particle formed by smaller particles through coagulation, usually promoted by chemical addition.
Hindered settling	Settling of particles in a liquid medium wherein particles do not behave as a single particles because of the interaction of other near field particles.
Solid Flows	Downward passage of solids in a thickening unit process.
Aerobic	Processes taking place in the presence of oxygen
Anaerobic	Processes taking place in the absence of oxygen
Mesophilic	Pertains to a group of microorganisms that thrive in a temperature range of about 30 to 40°C.
Thermophilic	Pertains to a group of microorganisms that thrive in a temperature range of about 40 to 50°C.

Digestion	A process where complex organic compounds are decomposed into methane and carbon dioxide gases by facilitative anaerobic microorganisms. Digestion can also be accomplished by aerobic bacteria, to produce carbon dioxide and ammonia
Pathogen	A microorganism that produces disease.
Disinfection	The elimination of pathogenic and other microorganisms by chemical addition or other means.
Tertiary treatment	Processes that are added to secondary waste treatment facilities to improve the quality of the effluent.

Abbreviation	Meaning
ASDA	New York State Atomic and Space Development Authority
AEC, USAEC	United States Atomic Energy Commission
Btu	British thermal unit (s)
BWR	Boiling water reactor
C_{steam}	Cost of steam, cents per 10^6 Btu
10 CFR 100	Title 10, U.S. Code of Federal Regulations, Part 100
COD	Chemical oxygen demand, ppm
CW	Cooling water
DEC	New York State Department of Environmental Conservation
ENR	Engineering News Record Construction Cost Index
EPA	United States Environmental Protection Agency
gpcd	Gallons per capita per day
gpd	Gallons per day
gpm	Gallons per minute
"Hg, "Hg abs	Absolute steam pressure, inches of mercury
HP	High pressure (steam), > 100 psia
HTGR	High temperature gas-cooled reactor

Kw, Kwe	Electrical kilowatts
LILCO	Long Island Lighting Company
LWR	Light water reactor
MBtu	Million Btu
MGD	Millions of gallons per day
MP	Medium pressure (steam, > 50 psia)
MSF	Multistage flash distillation system
Mw, Mwe	Electrical megawatts
Mwt	Thermal megawatts
NSSS	Nuclear steam supply system
O&M	Operations and maintenance (costs)
QC	Quality control
PHS, USPHS	United States Public Health Service
psia, psi (abs)	Absolute pressure, pounds per square inch
PWR	Pressurized water reactor
VTE	Vertical tube evaporator

APPENDIX A

TEMPERATURE EFFECTS ON WASTEWATER TREATMENT PROCESSES

CONTENTS

FIGURES

FIGURES
(continued)

FIGURES
(continued)

Number		Page

FIGURES
(continued)

SUMMARY OF FINDINGS

The effects of heat enrichment on the efficiency of common waste treatment unit operations were investigated based upon a review of generally available literature. The table below presents a summary of the units investigated and the change in size and in efficiency for these operations when operated at elevated temperatures. The basis of this table is an increase of temperature of from 20°C to 30°C.

SUMMARY OF UNITS SIZE AND EFFICIENCY
CHANGES DUE TO 10°C TEMPERATURE INCREASE
(20°C to 30°C)

Unit Operation	Change Due to Temperature Increase: % Change in Unit Size	Change Due to Temperature Increase: % Change in Unit Efficiency
Grit Chamber	16.5	13
Primary Clarifier	20	13
Aeration Basin	10	12
Trickling Filter	68	30
Stabilization Pond	48	15
Aerated Basin	52	8
Final Clarifier	20	-
Chlorine Contact Tank	28	-
Thickener	20.5	-
Anaerobic Digestor	38.5	8.2
Vacuum Filter	14	9
Centrifuge	20	25
Filtration (Strat.)	19	20
Rapid Sand		60
Backwash Rate		-16
Activated Carbon	-8	29
Foam Separation		-5
Nitrification	27	65
Denitrification (A.S.)	92	47
Ammonia Stripping	50	14
Anaerobic Column (Nitrogen Removal)		28
Pure O_2 Activated Sludge		114
Coagulation		50

Note: *Minus sign indicates an increase in unit size or a decrease in efficiency. No sign indicates a decrease in unit size or an increase in efficiency.*

SECTION A-I

INTRODUCTION

The effect of heat addition on the unit operations commonly found in a waste treatment facility has been investigated and is reported in this appendix. These investigations are based on a review of generally available literature. The unit operations evaluated include the following major operations:

1. Sedimentation
2. Solids thickening
3. Anaerobic digestion of solids
4. Anaerobic reduction of carbon and nitrogen
5. Aerobic bio-oxidation of carbon
6. Aerobic bio-oxidation of ammonia
7. Anaerobic bio-reduction of nitrate
8. Solids dewatering
9. Thermal processing of solids
10. Solids drying
11. Effluent disinfection

For discussion purposes, the unit operations are divided into the major categories of:

- Physical-chemical processes
- Sludge handling processes
- Biological processes
- Disinfection
- Advanced treatment processes

It is recognized that assignment of some processes in a specific category is somewhat arbitrary.

Graphical presentations showing possible process loadings and/or efficiency variations as a function of temperature were developed where possible. The sources of the correlations are presented in this appendix. These illustrations do not firmly establish process design criteria but, rather, qualitatively describe the effect of temperature on process performance. For most of the unit operations evaluated, further laboratory-scale or pilot-scale investigations would be required to firmly establish design relationships.

The sewage temperature at the inlet to the sewage treatment plant is influenced by many factors such as ambient temperature, length of interceptors, and source of the wastewater. Though variations about a mean value will occur throughout the annual cycle, a temperature of 20°C is selected as a basis for comparing the temperature effects. Most of the illustrations show values relative to 20°C.

SECTION A-II

PHYSICAL-CHEMICAL PROCESSES

Increasing the temperature has a significant effect upon the physical-chemical processes that are commonly employed in waste treatment. The physical processes – grit removal, clarification, thickening, sludge dewatering, and flotation – involve a separation of solid material from water. Since this physical separation process is mainly a function of the fluid viscosity, decreasing the viscosity of a fluid by increasing the temperature can increase the efficiency of a separation process.

Chemical reaction rates are enhanced by elevated temperatures. Chemicals are used in wastewater treatment plants primarily to aid in the liquid-solid separation processes. However, certain processes (disinfection, phosphorus removal, and biological denitrification) employ chemicals for a specific function other than as a separation aid. These processes will be discussed in later sections of this appendix. This section will discuss the physical-chemical processes that primarily effect a liquid-solids separation.

Grit Removal

Grit removal in large wastewater treatment plants is generally accomplished in grit chambers which are designed to separate inert solids from the flowing medium. These inert solids are removed to alleviate excessive wear on the mechanical equipment (pumps) and to minimize possible interference with wastewater treatment processes following the grit chambers.

When a discrete particle settles in a quiescent medium, the particle will accelerate until the frictional resistance or drag force equals the gravitational force. Thereafter, the particle will settle at a constant speed. For spherical particles, the terminal settling velocity is as follows [5, 6]:

$$V_s = \left[\frac{4}{3}\frac{g}{C_D}(S_s-1)d\right]^{1/2} \quad \text{..............(\#1)}$$

where: V_s = Terminal settling velocity

g = Gravitational constant

C_D = Drag coefficient

S_s = Specific gravity

d = Diameter of sphere

The drag coefficient (C_D) is a function of the Reynolds number. For a Reynolds number $(R) < 10^4$, the value of C_D is as follows:

$$C_D = \frac{24}{R} + \frac{3}{\sqrt{R}} + 0.34 \quad \text{.............. (\#2)}$$

For a Reynolds number less than 0.5, the drag coefficient can be described as:

$$C_D = \frac{24}{R} \quad \text{.............................. (\#3)}$$

Equation (#1) similarly reduces to the following when the Reynolds number is less than 0.5:

$$V_s = \frac{g}{18} \frac{(S_s - 1)}{\nu} d^2 \quad \text{................... (\#4)}$$

where: ν = Coefficient of kinematic viscosity

This relationship (#4), known as Stoke's law, describes the terminal settling velocity derived for spherical particles that settle discretely. Discrete sedimentation assumes that the particles settle without colliding or interacting with any other particles. From Equation (#4), the influence of viscosity and hence temperature is readily apparent.

Of the three unit operations which separate solids from the waste, namely, grit removal, primary clarification, and secondary clarification, the settling phenomenon in a grit chamber most closely agrees with discrete particle settling described by Equation (#1) or (#4). Grit chambers are generally designed to remove inert particles that have a size greater than 2×10^{-2} cm and a specific gravity of approximately 2.6.

For a continuous flow tank with turbulence, the works of Hazen and/or Dobbins can be used to compute solids removal efficiencies. Hazen's real tank theory [6] states that the efficiency of a settling basin is a function of the particle settling velocity, the surface area of the basin, rate of flow, and the hydraulic characteristics of the basin. The Hazen relationship is:

$$\text{Removal of Suspended Solids} = 1 - \left[1 + \frac{nVs}{Q/A}\right]^{-1/n} \quad \ldots(\#5)$$

where: n = Coefficient that identifies basin performance

n = 1/3 for good performance

n = 1/2 for poor performance

V_s = Terminal settling velocity

Q = Hydraulic flow rate

A = Surface area

The theoretical increase in removal efficiency and decrease in grit chamber length with rising temperatures have been computed, using Hazen's model (#5), as shown on Figures 1 and 2. Figure 1 shows the relative suspended solids remaining at various temperatures. A base removal of 75% at 20°C in a good performance chamber is used. At 30°C, approximately 20% of the suspended solids remaining at 20°C would be removed. This corresponds to an overall suspended solids removal of approximately 80% in a grit chamber sized for 75% removal at 20°C.

Figure 2 illustrates the relative change in grit chamber size with variation in temperature and constant removal efficiency. The width of the grit chamber is kept constant and a chamber with good performance characteristics is used. At 30°C, grit chamber length can be reduced by approximately 20% over an equivalent chamber designed at 20°C.

The composition of settleable and non-settleable solids in the wastewater is an important factor in analyzing the performance of grit chambers. Some experimental data show that the actual removal efficiencies in grit chambers are lower than those predicted from Hazen's real tank theory [6]. The actual performance of a grit chamber and other physical separation units is therefore very much a function of raw waste characteristics.

GRIT CHAMBER

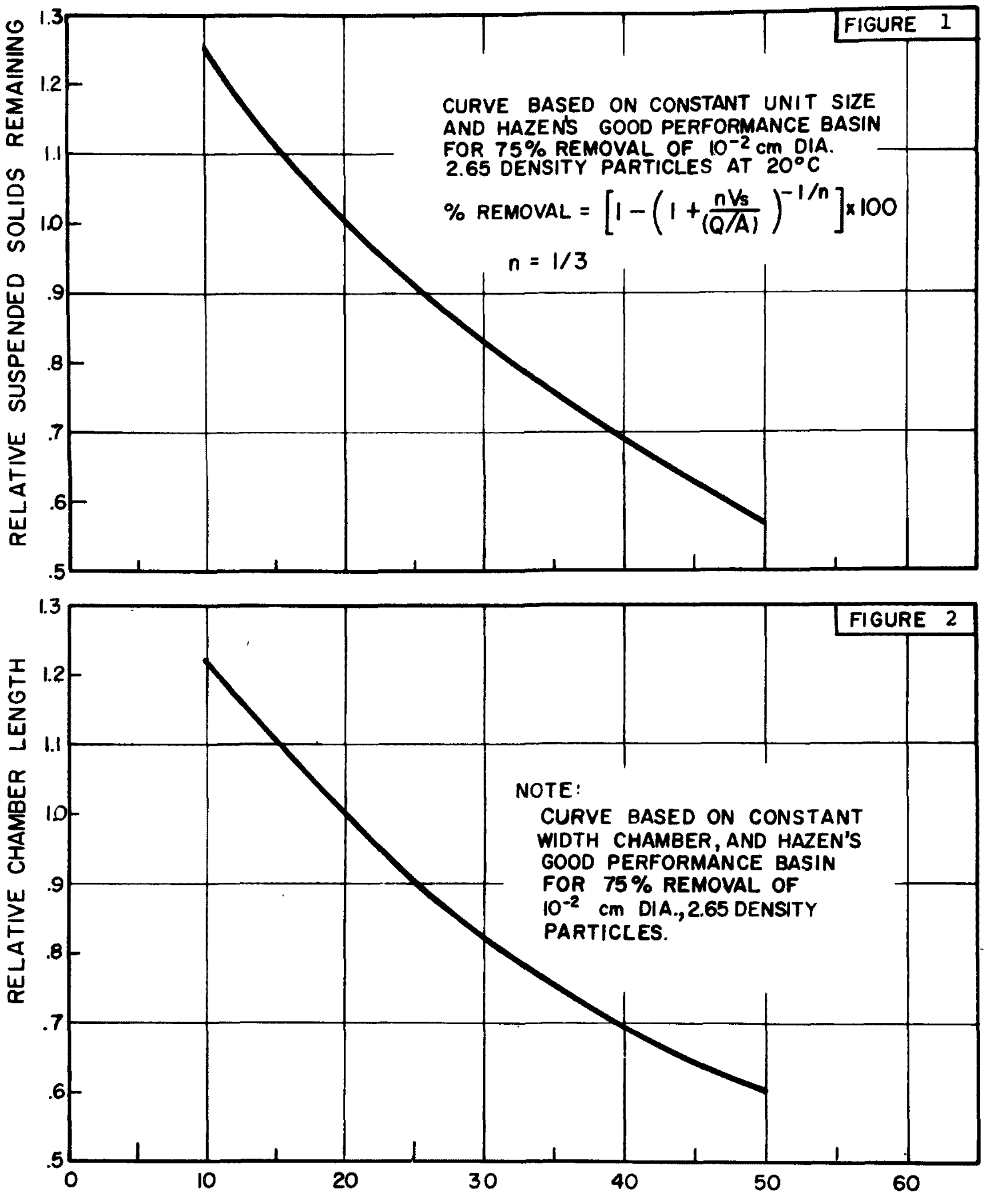


FIGURE 1 EFFECT OF TEMPERATURE ON THE EFFICIENCY OF GRIT REMOVAL, UNIT SIZE HELD CONSTANT.

FIGURE 2 EFFECT OF TEMPERATURE ON GRIT CHAMBER SIZE TO ACHIEVE A CONSTANT 75% GRIT REMOVAL EFFICIENCY.

Sedimentation

Generally, the gross settleable solids present in wastewater are removed by gravity settling in primary sedimentation basins. This process also provides a reduction of the waste load to subsequent treatment units. The settling phenomenon in primary basins can be described as a mixture of discrete and flocculent particle sedimentation. The particles removed are of smaller size and have a lower specific gravity than the grit particles removed in the grit chambers.

As an illustration of the theoretical effect of temperature on the settling of discrete particles in a sedimentation basin, Figures 3 and 4 are presented. These curves apply Hazen's real tank model in rectangular basins operating at an overflow rate of 800 gpd/SF. A suspended solids removal efficiency of 50% at 20°C is utilized for these "good performance" basins.

Increase in wastewater temperature from 20°C to 30°C will result in an increase in suspended solids removal from 50% to 56% in a constant size sedimentation basin. To achieve an equivalent suspended solids removal of 50% at the elevated temperature, a basin approximately 20% smaller than would be required at 20°C is needed. These changes in tank size and efficiency are based on theoretical considerations of discrete particle settling.

For flocculent particles, the settling velocity in a sedimentation basin is variable. During the settling process, the particles coalesce in the basin, thereby affecting the size and density of the aggregate. The net effect is an increase in velocity as the particles collide. Since the settling velocity is not constant and depends upon the flocculent nature of the solids, the removal efficiency depends not only on the surface area and flow rate, but also on the detention time. As the temperature increases, the settling velocity of the flocculated solids increases. Additionally, temperature is believed to have a beneficial effect on the flocculating characteristics of the solids [8]. Pilot-scale testing of flocculent settling at various temperatures would provide information on removal efficiencies to be expected in secondary clarifiers.

Many factors reduce the efficiency of a prototype sedimentation basin. Various currents such as wind-induced surface currents, convection currents, density currents,

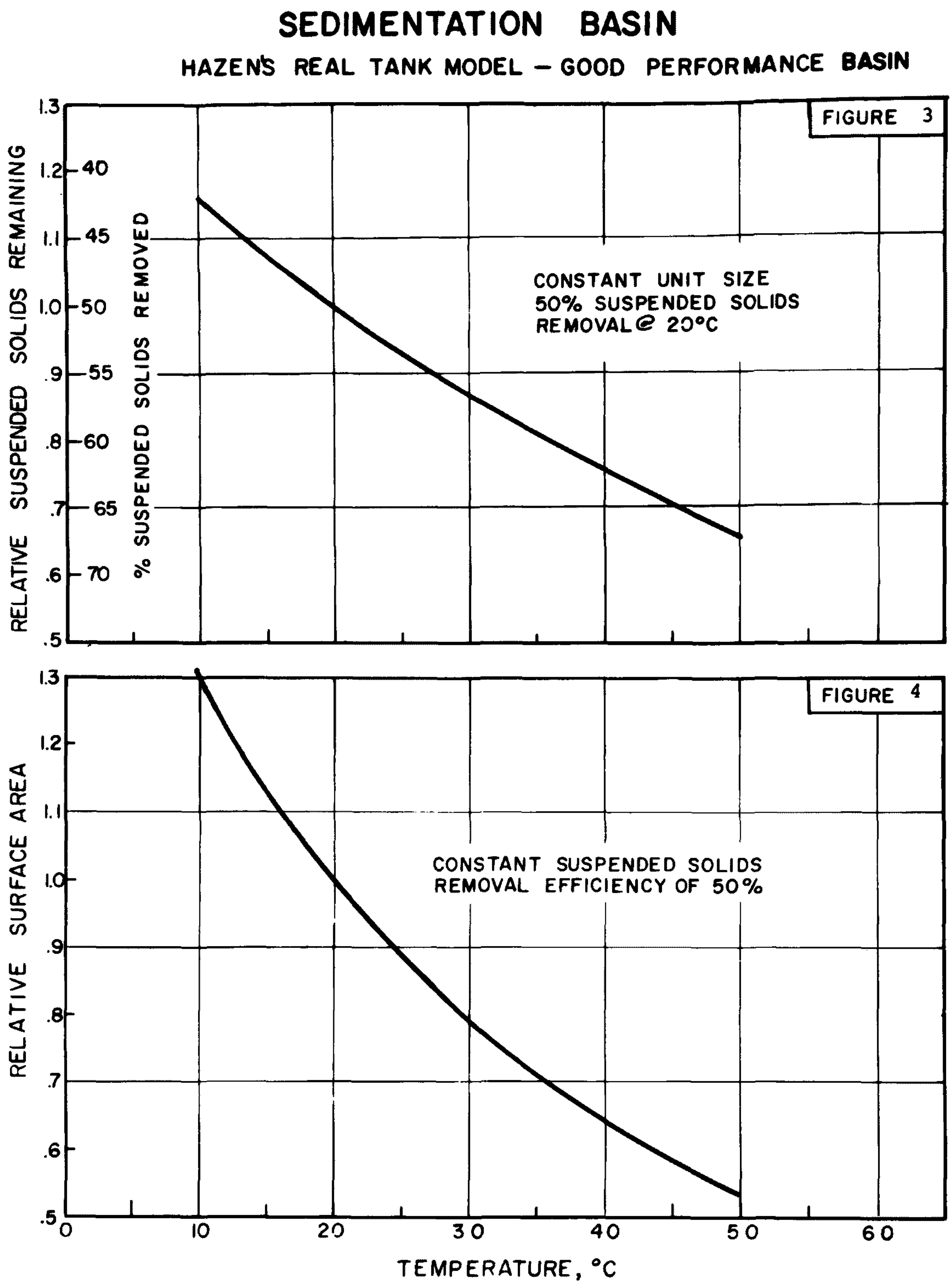


FIGURE 3 EFFECT OF TEMPERATURE ON THE SUSPENDED SOLIDS REMOVAL IN SEDIMENTATION BASINS, UNIT SIZE HELD CONSTANT.

FIGURE 4 EFFECT OF TEMPERATURE ON SEDIMENTATION BASIN SURFACE AREA TO ACHIEVE A CONSTANT 50% SUSPENDED SOLIDS REMOVAL.

and velocity distribution tend to contribute to short-circuiting of flow through the basin. Thermal gradients may exist within an uncovered basin. These gradients can have a deleterious effect on the basin flow pattern and removals. The effects of thermally induced gradients have to be evaluated on at least a pilot scale. For the purposes of analysis, the reductions in size presented on Figure 4 will be utilized for both primary and secondary clarifiers.

Coagulation

A wide variety of organic and inorganic solids in wastewater will not be removed by sedimentation unless agglomerated into larger particles. These particles are stabilized or kept separate by electrical and physical forces. Since natural destabilization forces are generally not sufficient to allow efficient solids removal, chemical coagulants are employed to stabilize and agglomerate the solids. For each combination of coagulant and wastewater, there is an optimum dosage of coagulant and an optimum pH range for coagulation.

Little experimental information is available on the effect of temperature on coagulation [6]. Velz [10] cites detrimental effects of higher temperatures on coagulation and removal of color. Higher alum dosages were required at elevated temperatures to achieve an equivalent effluent color concentration. He also related temperature, dosage, and time of appearance of first floc, all illustrating the disadvantages of high-temperature coagulation. However, these studies were performed at constantly changing pH values which were considered of minor importance at that time.

Camp [11] found that coagulant dosage and temperature could change optimum pH values markedly. With pH adjustment, Camp found shorter settling times at higher temperatures. Renn [8] also found shorter floc formation times at higher temperatures and that the floc was generally found to be coarser in nature. Willcomb [12] explains that, as a result of the increased viscosity of the water and the surface tension change in the floc at low temperatures, coalescing tendencies are resisted. Prolonged agitation would then be necessary to enlarge floc sizes to settleable proportions. Parsons [13] presented ranges of removal efficiency for coagulation of domestic sewage. These efficiencies range from 65% to 85% for removal of suspended solids and from 45% to 75% for removal of BOD.

Coagulation of effluent from secondary biological treatment facilities has also been evaluated. Stukenberg [14] achieved better BOD and COD removals at warmer operational temperatures.

The optimum conditions for flocculation are determined by three variables: the chemical dose, the pH, and the temperature. Renn [8] determined that the isoelectric point, or optimum pH value for coagulation, varies with temperature. It has been found that, at the optimum pH value, the required coagulant dosage decreases as the temperature increases. Additionally, the time of floc formation decreases as temperature increases.

As an illustrative example of the data reported on the effect of heat on coagulation, Figures 5 and 6 are presented. The data presented are taken from studies performed on municipal water supplies. Temperatures evaluated in these studies range from approximately 5°C to 28°C. Figure 5 indicates a rapid increase in required alum dosage to achieve satisfactory flocculation at temperatures lower than 10°C.

Figure 6 presents the time required to form a good floc at optimum pH and with varying chemical dosage. The figure reveals that chemical dosage is the most significant parameter affecting floc formation in the temperature range studied. At a chemical dosage of 2 ppm at 20°C, approximately 20 minutes were required for good floc formation. This floc formation time decreased to about 25 minutes at 28°C with the same chemical dosage. However, at a chemical dosage of 4.3 ppm, floc formation times remained constant.

Studies of coagulation at temperatures above 28°C were not found. Additionally, those studies that have been reviewed do not indicate the quantity or cost of pH adjustment. Wet testing on specific wastewaters is necessary.

Gravity Filtration

With the exception of gravity sedimentation, deep bed filtration is the most widely used unit process for liquid-solids separation. Recently, it has been employed in physical-chemical systems for polishing effluent prior to discharge [9].

The temperature at which filtration takes place has a large influence on the process. Filter bed variables affected by temperature are:

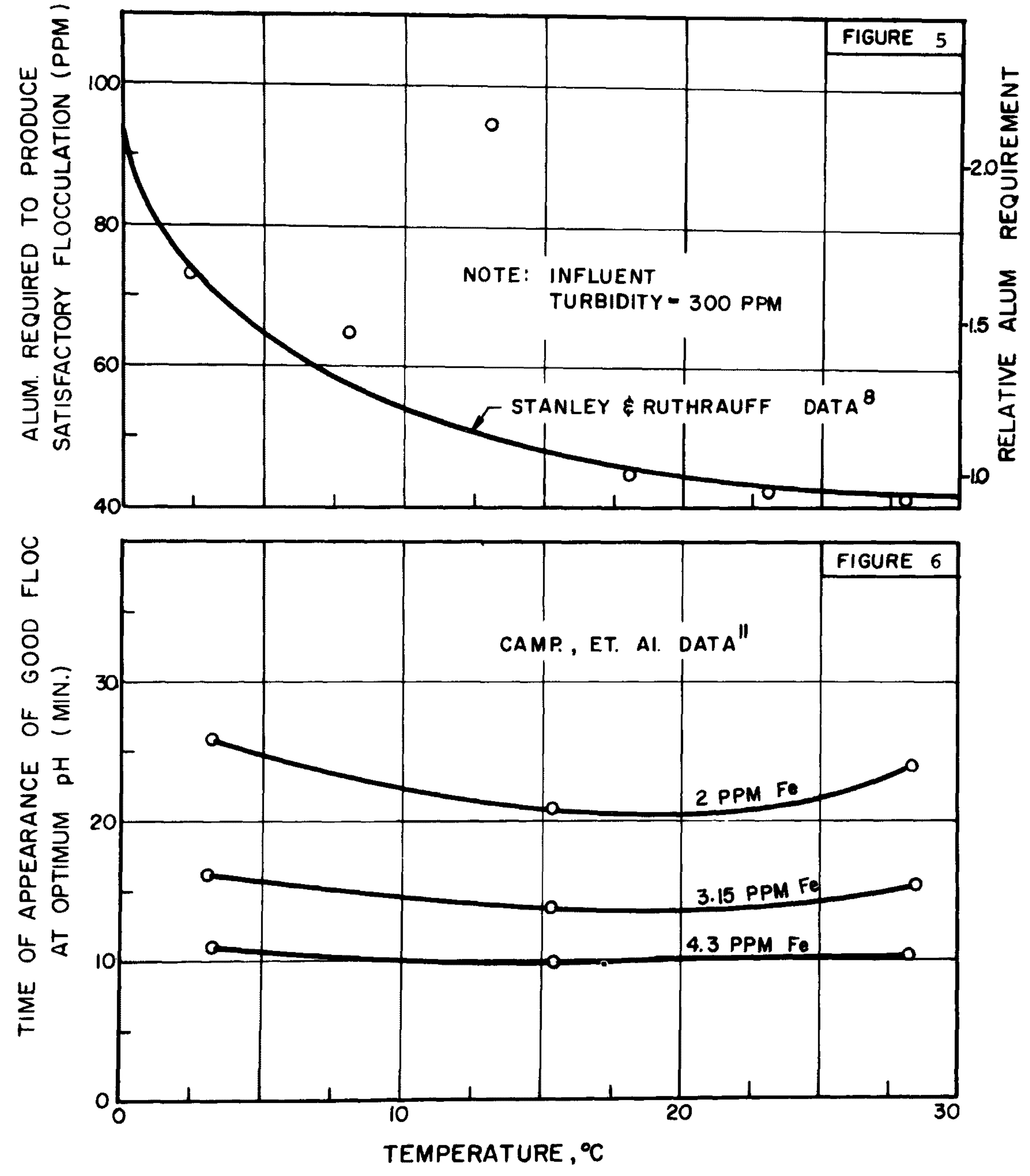


FIGURE 5 EFFECT OF TEMPERATURE ON THE ALUM DOSAGE REQUIRED TO PRODUCE SATISFACTORY FLOCCULATION.

FIGURE 6 EFFECT OF TEMPERATURE ON THE TIME REQUIRED FOR FLOC FORMATION.

(a) Depth of media
(b) Rate of filtration
(c) Expansion of bed on backwash
(d) Efficiency of filtration
(e) Head losses

Head losses through the filter are directly proportional to viscosity and, therefore, are reduced by temperature increase.

A relationship for head loss in a clean stratified bed sand filter was developed by Kozeny and modified by Fair and Hatch [6] to:

$$\frac{h}{l} = \frac{k}{g} \nu v \frac{(1-f)^2}{f^3} \left(\frac{6}{\Psi}\right)^2 \sum_{i=1}^{n} \frac{Pi}{di^2} \quad \ldots\ldots\ldots(\#6)$$

where: $\frac{h}{l}$ = Head loss in ft/ft length

k = Coefficient of permeability

ν = Kinematic viscosity

v = Liquid velocity through filter

f = Porosity of filter media

Ψ = Spheroscity of filter media

Pi = Fraction analyzed sand

di = Average diameter of sieved sand

Based on this equation, an illustrative example of the theoretical effect of temperature on a clean unstratified sand filter is presented on Figure 7. An increase in water temperature from 20°C to 30°C will result in a decrease in head loss of approximately 20%. A temperature increase to 60°C will result in a 50% reduction in head loss through the sand bed. The same equation (#6) can be used to compute the change in filter area required for a constant head loss and flow rate. Figure 8 presents the relative capacity of a sand filter with constant head and variable temperature. An increase of water temperature from 20°C to 30°C will require a sand filter of approximately 20% less area or depth than at 20°C.

GRAVITY FILTRATION — STRATIFIED BED

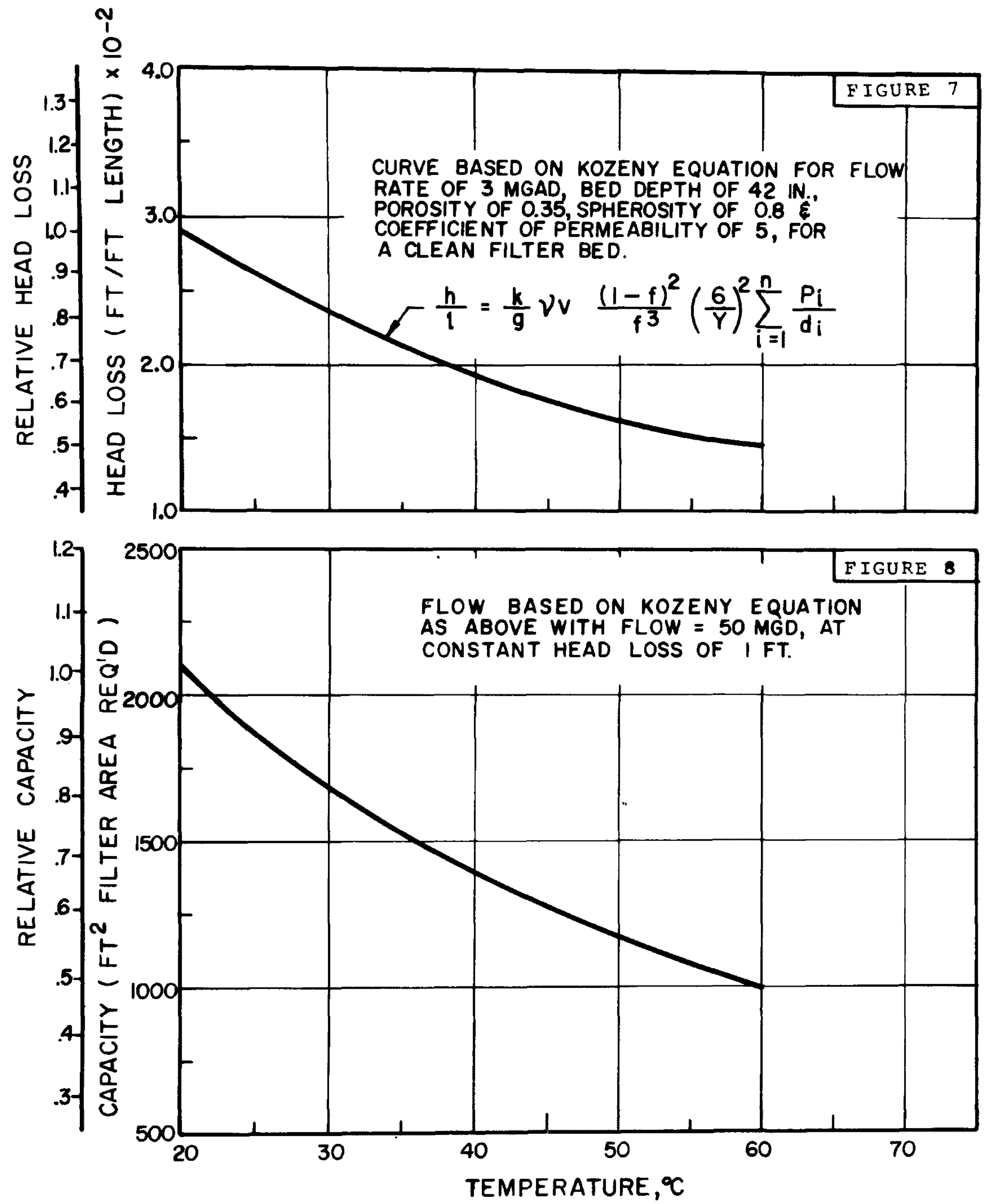


FIGURE 7 EFFECT OF TEMPERATURE ON HEAD LOSS THROUGH STRATIFIED BED GRAVITY FILTERS.

FIGURE 8 EFFECT OF TEMPERATURE ON FILTER CAPACITY TO PRODUCE A CONSTANT HEAD LOSS IN STRATIFIED BED GRAVITY FILTERS.

The backwash water rate required to clean a sand filter has been found to vary with temperature [16]. A relationship, based on Hazen's formula, for backwash rate is:

$$R = 30d^{1.5}\ (1 + 0.060x)\ \frac{(t + 30)}{80} \quad .....(\#7)$$

where: R = Backwash rate (in./min)

d = Effective sand size (mm)

x = % bed expansion (expressed as a whole number)

t = Wash water temperature (°F)

An illustrative example of the theoretical variation in backwash rate with temperature is presented on Figure 9. Also indicated on this figure are the limited data reported by Lawrence [16]. The portion of the curve above 25°C is shown as a dashed line to show that it is an extrapolation of the basic equation. A temperature increase from 20°C to 30°C will result in a requirement of 20% higher backwash rate for an equal bed expansion.

The spatial distribution of solids or turbidity was experimentally determined by Ives [15] to be described by the following:

$$-\frac{\partial c}{\partial l} = \lambda c \quad ....(\#7a)$$

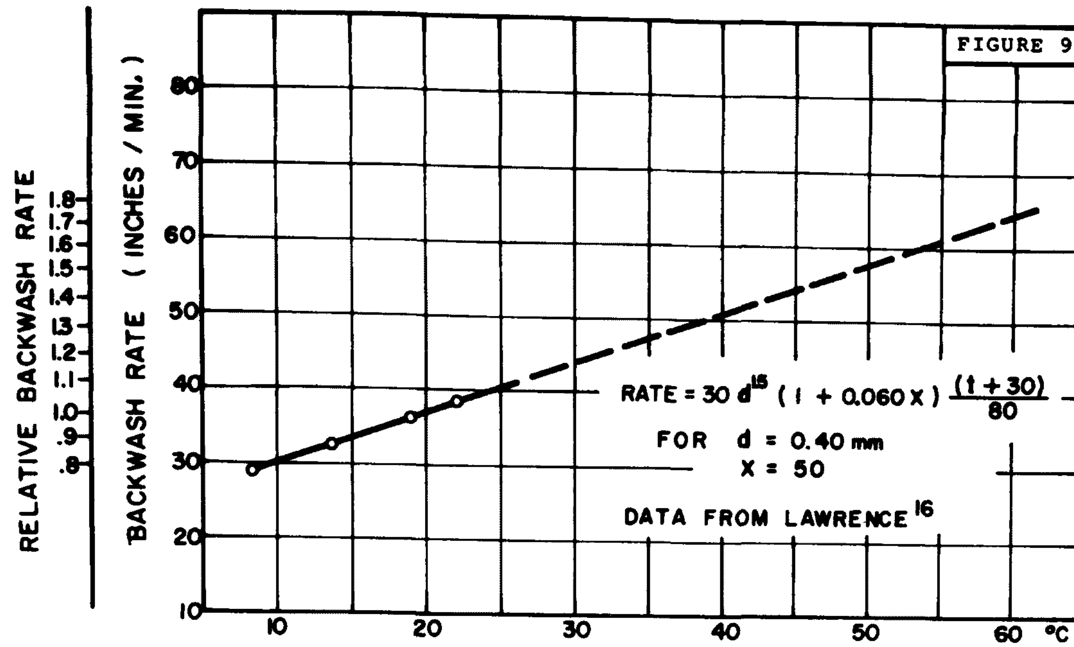
where: c = Concentration of suspension

l = Depth of filter medium

λ = Filter coefficient

The filter coefficient (λ) is a function of the amount of solids that are deposited in the filter. The coefficient (λ) in Equation (#7a) is obtained for a clean filter operating with an initial filter coefficient of λ_0. The equation which describes the variation of the filter coefficient is as follows [15]:

$$\lambda = \lambda_0 + c\sigma - \frac{\phi\sigma^2}{f-\sigma} \quad ....(\#8)$$



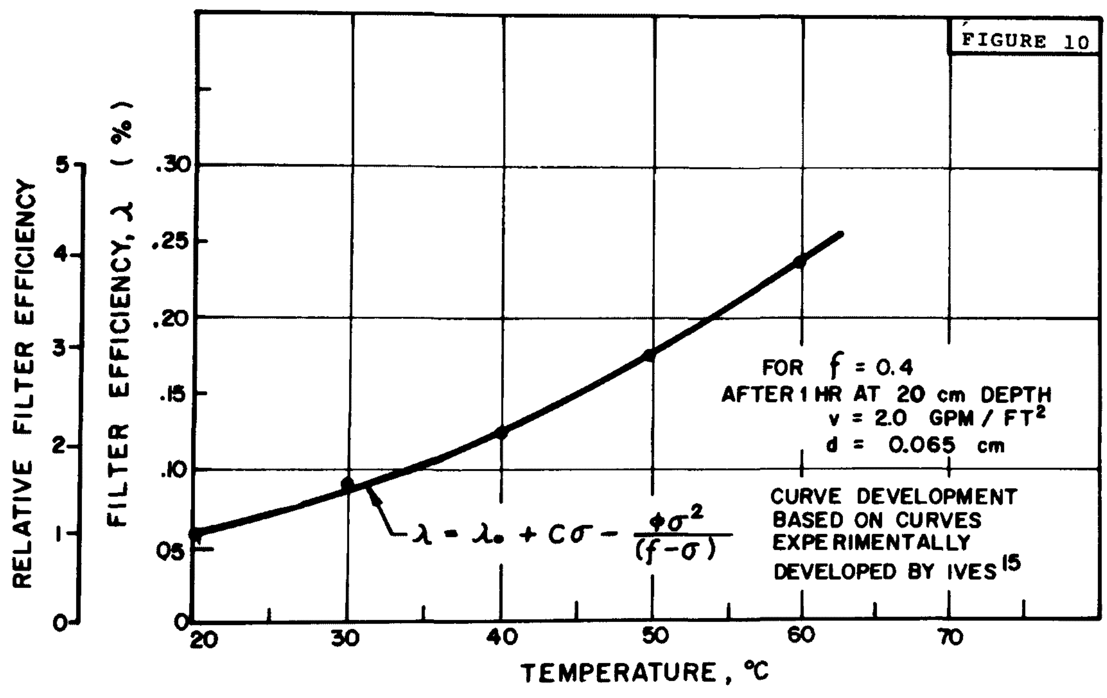


FIGURE 9 EFFECT OF TEMPERATURE ON REQUIRED BACKWASH RATE FOR GRAVITY FILTERS.

FIGURE 10 EFFECT OF TEMPERATURE ON FILTER EFFICIENCY, CONSTANT FILTER SIZE.

where: λ_0 = Initial (clean) filter coefficient

σ = Volume of deposit per unit filter volume

ϕ = A second filter constant

f = Porosity of filter medium

The filter coefficient is a direct measure of the filter efficiency. An illustrative example of the theoretical effect of elevated temperature on filter efficiency is presented on Figure 10. For this figure, λ_0, c and ϕ were determined at various temperatures using data presented by Ives [15]. For the conditions cited on Figure 10, λ was then computed as shown. A temperature increase from 20°C to 30°C will result in a 50% increase in the filter coefficient and therefore the filter efficiency. Full-scale evaluation of the effects of heat addition to sand filters has to be performed for temperatures above 25°C to verify the theoretical relationship presented in this discussion.

Flotation

Flotation is generally confined to a process in which air is dissolved into the wastewater. The mixture is held under pressure to ensure adequate solution of the air into the liquid. The mixture is then released to atmospheric pressure upon which the air in the form of small bubbles is released. These bubbles either become enmeshed in or attached to the suspended material.

The performance of the unit depends upon having sufficient air bubbles to float the suspended material. Effluent quality and/or solids concentration in the float are related to an air/solids ratio which is defined as the pounds of air released per pound of initial suspended solids [17]. The rise rate or velocity of the solids is also directly influenced by the air/solids ratio. Parameters important for flotation processes include: (1) pressure, (2) solids concentration, (3) detention time, (4) type and quality of waste and sludge and its volatile content, (5) solids and hydraulic loading rates, (6) temperature, (7) recycle ratio, (8) air to solids ratio, and (9) use of chemical aids [18].

Flotation processes include: (1) dispersed air-flotation, (2) dissolved air-vacuum flotation, (3) dissolved air-pressure flotation, (4) biological flotation. In application, the first two methods may be employed for wastewater treatment while the last two are for sludge thickening operations.

At higher wastewater temperatures, the decrease in liquid viscosity will lessen the resistance to liquid-solid separation. However, the solubility of air in water is inversely proportional to the temperature and, to maintain comparable air/solids ratios at higher temperatures, the pressure will have to be correspondingly increased. The actual effect of elevated temperature upon sizing a specific unit would have to be determined in laboratory and pilot studies.

Sludge thickening by heat flotation has been investigated by Malina [20] and Laboon [21]. In these processes, as the sludges are heated the solid material tends to rise as a mat, resulting from a lifting effect caused by the release of absorbed gases in the form of small bubbles. The application of these processes depends upon the amount of entrained gases in the sludge, the temperature to which the sludge is raised, the point of heat addition, and the quantity of sludge to be thickened.

Because many primary variables exist for flotation processes as indicated, and because some of these variables are interrelated or affected by prior processes in a treatment system, when a flotation unit is selected for operation at high temperature, pilot plant studies are suggested for effective design of the unit.

SECTION A-III

SLUDGE HANDLING PROCESSES

Sludge handling processes have been investigated and reported in great depth by Burd [18]. Some 450 references are cited in his work. For the current study, we have limited our investigation to the major unit processes. The reader is directed to the Burd report for discussion of specific unit processes not present in this report. The material which follows has been arranged in two categories: (1) thickening and (2) dewatering. Generally, the conditioning and dewatering unit operations respond to temperature increases in ways which incorporate many of the previously developed temperature relationships. Where this is the case, general reference will be made to the previous development.

Thickening

Primary sludge and secondary sludge (waste-activated sludge) require further concentration prior to digestion or dewatering. A gravity-type or flotation-type thickener is generally employed for this application. The solids settle or rise at different velocities depending upon the solids concentration and temperature. Previous discussions have dealt with Stoke's law and the change in discrete particle settling velocity with increases in temperature (see Grit Removal).

A relationship has been developed to describe the hindered settling velocity of flocculent sludges as a function of sludge characteristics, concentration, and the discrete particle settling velocity [32]. The relationship is described by the following equation:

$$V_i = V_o (1 - C)^n \quad \text{.....................(\#9)}$$

where: V_i = Critical settling velocity of the sludge at concentration C

V_o = Discrete settling velocity of a particle

n = Empirical value depending upon the sludge characteristic

C = Concentration of solids at velocity V_o

Solids are transported to the bottom of a gravity thickener by two mechanisms: their subsidence due to gravity and the bulk downward transport due to sludge withdrawal from the bottom of the thickener. The rate at which solids of concentration C_i pass downward in the thickener can be described by the following [32]:

$$G = C_i V_i + C_i U \quad \text{.......................(\#10)}$$

where: G = Solids flux, expressed in lbs/day-SF

V_i = Settling velocity of the sludge at concentration C_i

U = Average downward velocity caused by removal of the sludge from the bottom of the tank

The term C_iV_i depends upon the settling characteristics of the sludge, while C_iU is a variable controlled by the operation. Characteristically, the solids flux achieves a minimum value which provides the basis of a design to ensure sufficient thickener surface area to meet the area requirements of this minimum solids flux.

A more familiar form of this relationship is presented by Eckenfelder [25], based on a material balance between the influent, the underflow, and the effluent, as:

$$U.A. = \frac{(1/C_i - 1/C_u)}{V_i} \quad \text{.................(\#11)}$$

where: U.A. = Unit area (SF/lb solids/day)

C_i = Solids concentration at settling velocity V_i lb/CF

C_u = Underflow solids concentration, lb/CF

The unit area is inversely proportional to the particle settling velocity, which is affected by viscosity (and therefore temperature) changes.

An illustrative example of the theoretical effect of temperature increase on thickener requirements is presented on

Figures 11 and 12. Figure 11 presents the relative mass loading (lbs/SF/day) as a function of temperature. The mass loading is the reciprocal of unit area as described in Equation (#11). Figure 11 describes the relative increase in allowable solids application rate to a thickener of known size operating at a constant underflow concentration. A temperature increase from 20°C to 30°C would allow an increase of 27% in the mass loading on a thickener for the same underflow solids concentration. Figure 12 presents the relative unit area required as a function of temperature. A temperature increase from 20°C to 30°C would permit a 21% reduction in the size required at 20°C to achieve the same underflow solids concentration.

The relationships presented on Figures 11 and 12 assume that the sludge settling characteristics [21] and sludge blanket behavior (gasification) would not change significantly by increases in temperature. Experimental evaluations are required to determine what these effects might be.

Data on the compaction characteristics of pure oxygen activated sludge have been presented by Stamberg [125]. These data are presented on Figures 13 and 14. Figure 13 presents the change in initial batch flux (lb/SF/day) with initial mixed liquor concentration and temperature. At 4,000 mg/l MLSS, an increase in temperature from 10°C to 29°C resulted in a 64% increase in the flux rate (55 lb/SF/day to 90 lb/SF/day). Figure 12 presents the change in initial settling velocity (ft/hr) with initial mixed liquor concentration and temperature. At 6,000 mg/l MLSS, an increase in temperature from 23°C to 27°C resulted in a 71% increase in the initial settling velocity (7 ft/hr to 12 ft/hr).

Dewatering

The dewatering of waste sludges is a cumbersome and costly part of the wastewater treatment plant operation. The main objectives of these processes are to reduce the sludge volume for ease and economy of disposal. Commonly used sludge dewatering processes include (1) vacuum filtration, (2) pressure filtration, and (3) centrifugation. Discussion of the effect of elevated temperatures on these processes follows.

GRAVITY THICKENING

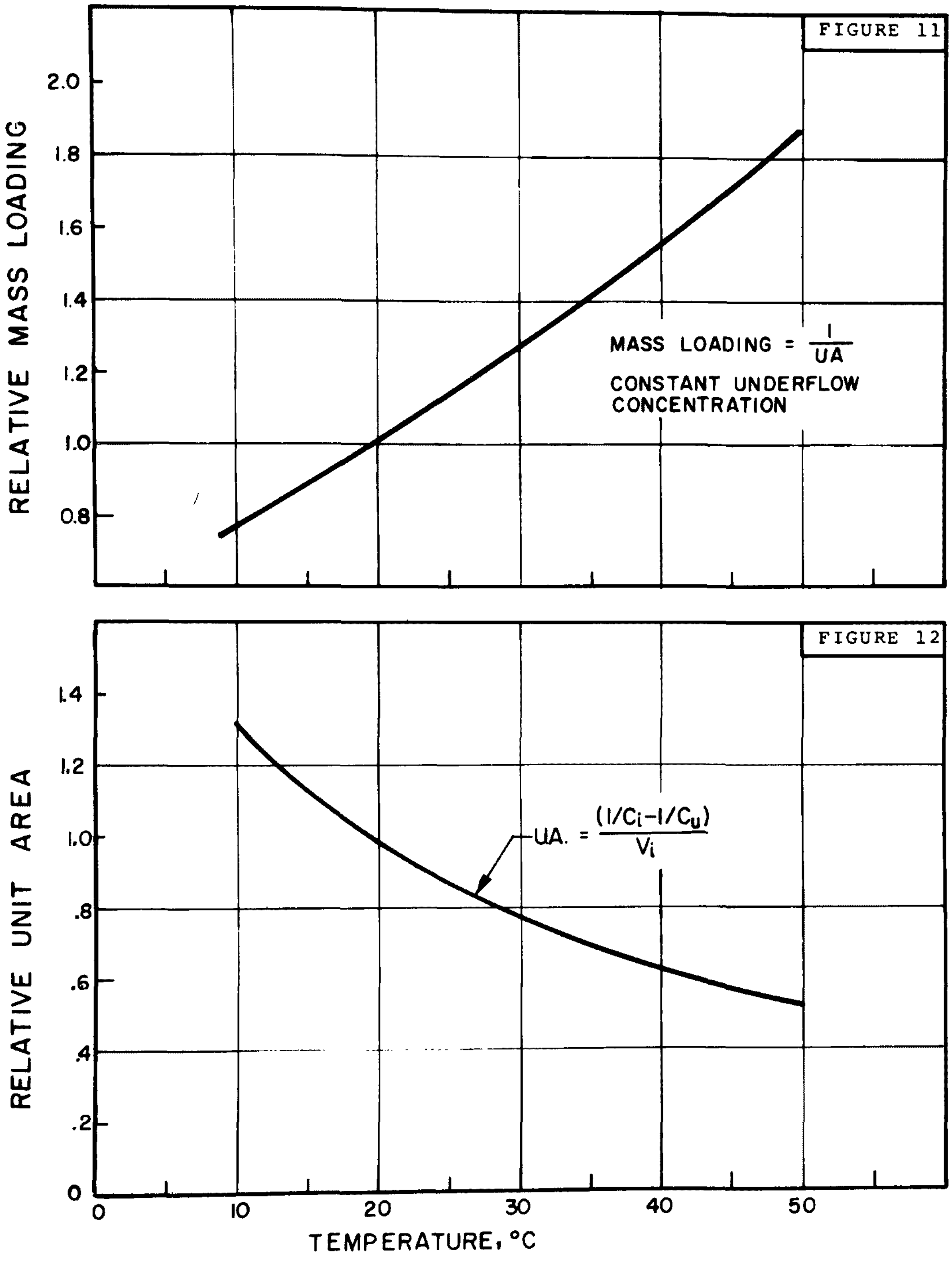


FIGURE 11 EFFECT OF TEMPERATURE ON MASS LOADING RATE TO THICKENER TO PRODUCE A CONSTANT UNDERFLOW CONCENTRATION.

FIGURE 12 EFFECT OF TEMPERATURE ON THICKENER SIZE TO PRODUCE A CONSTANT UNDERFLOW CONCENTRATION.

PURE OXYGEN ACTIVATED SLUDGE
COMPACTION CHARACTERISTICS

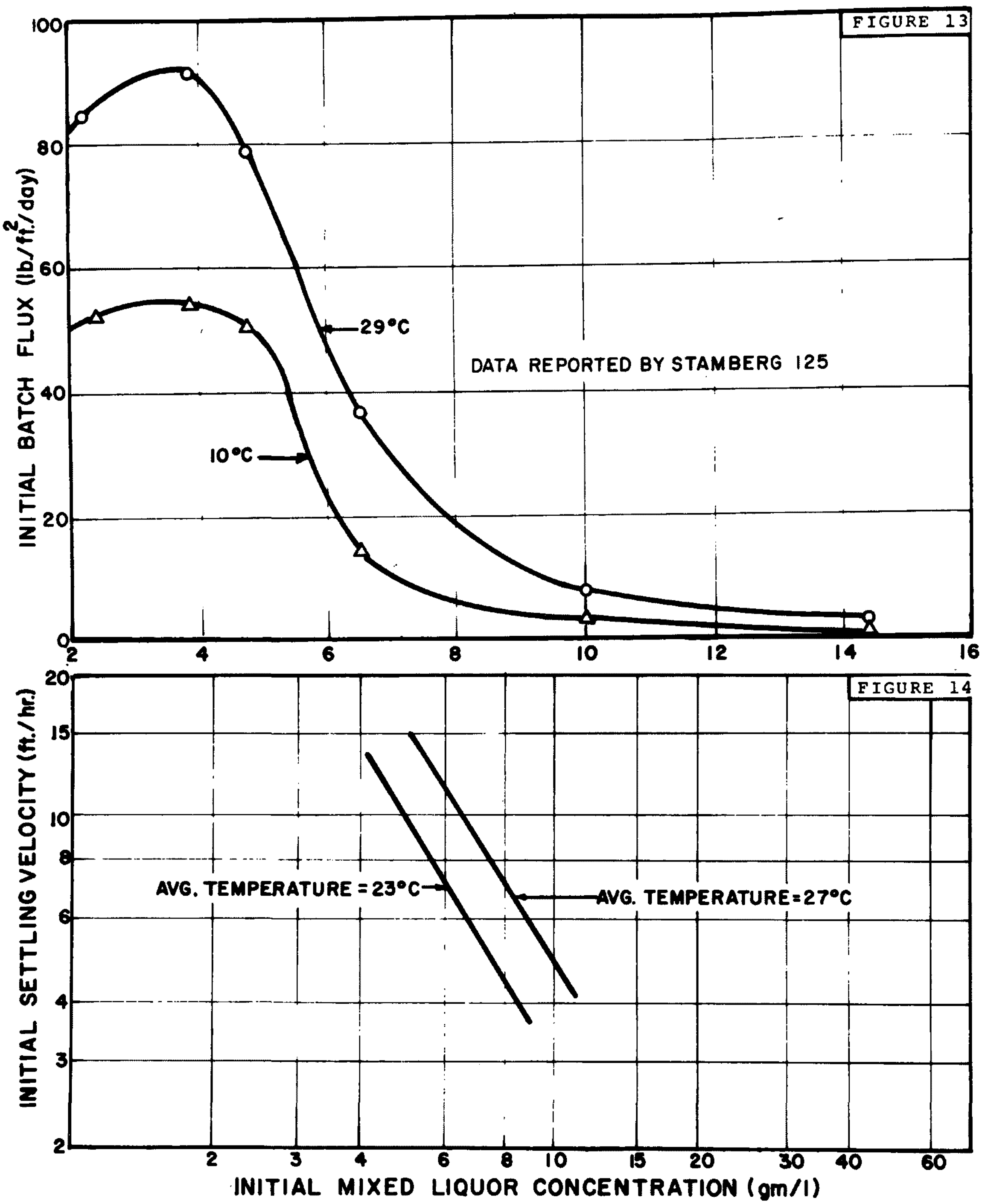


FIGURE 13 & 14 EFFECT OF TEMPERATURE ON THE COMPACTION CHARACTERISTICS OF SLUDGE PRODUCED BY THE PURE OXYGEN ACTIVATED SLUDGE PROCESS.

Vacuum Filtration

Vacuum filtration is commonly carried out on slowly rotating drum filters in a continuous operation. Variables affecting the performance of vacuum filters include: solids concentration, sludge type, temperature, sludge and filtrate (liquid) viscosity, and chemical composition.

The rate of filtration of sludges has been formulated according to Poiseville's and D'Arey's laws by Carmen and Coakley, as follows [25]:

$$\frac{dV}{dt} = \frac{PA^2}{\mu \left(rcV + R_m A\right)} \quad \text{....................(\#12)}$$

where: V = Volume of filtrate

t = Cycle time (approximates form time in continuous drum filters)

P = Vacuum

A = Filtration area

μ = Filtrate viscosity

r = Specific resistance

c = Weight of solids per unit volume of filtrate

R_m is the initial resistance of the filter medium and can usually be neglected as compared with the resistance developed by the filter cake. This equation can be modified to express filter loading rate (neglecting the initial resistance of the filter medium) as follows:

$$L = 35.7 \left(\frac{CP}{\mu Rt}\right)^{1/2} \quad \text{....................(\#13)}$$

where: $R = r \times 10^7 \ sec^2/gm$

P = Vacuum, psi

C = Solids deposited per unit volume filtrate, gm/ml

μ = Filtrate viscosity, centipoises

t = Form time, min

Since the filter loading rate is inversely proportional to the square root of the viscosity, the loading rate increases with increasing temperature. The specific resistance will similarly decrease at elevated temperatures. However, this effect has not been quantified since the reduction in specific resistance depends upon the sludge characteristics.

The theoretical effect of elevated temperatures can be estimated using Equation (#13). Figure 15 presents the theoretical filter yield as a function of temperature. Figure 16 presents the relative filter area required for a constant loading rate as a function of temperature. For a 10°C temperature rise from 20°C to 30°C, the filter area required drops 10%.

Pressure Filtration

Like vacuum filtration, a porous medium is used in leaf filters to separate solids from liquids [18]. (Leaf filters are the most commonly used pressure filter.) As sludges are forced onto the medium under pressure, the solids are captured in the medium pores and build up on the medium surface.

In general, when the liquid phase is highly viscous, or when the solids are so fine that vacuum filtration is too slow, pressure filtration provides a convenient solution to the separation problem.

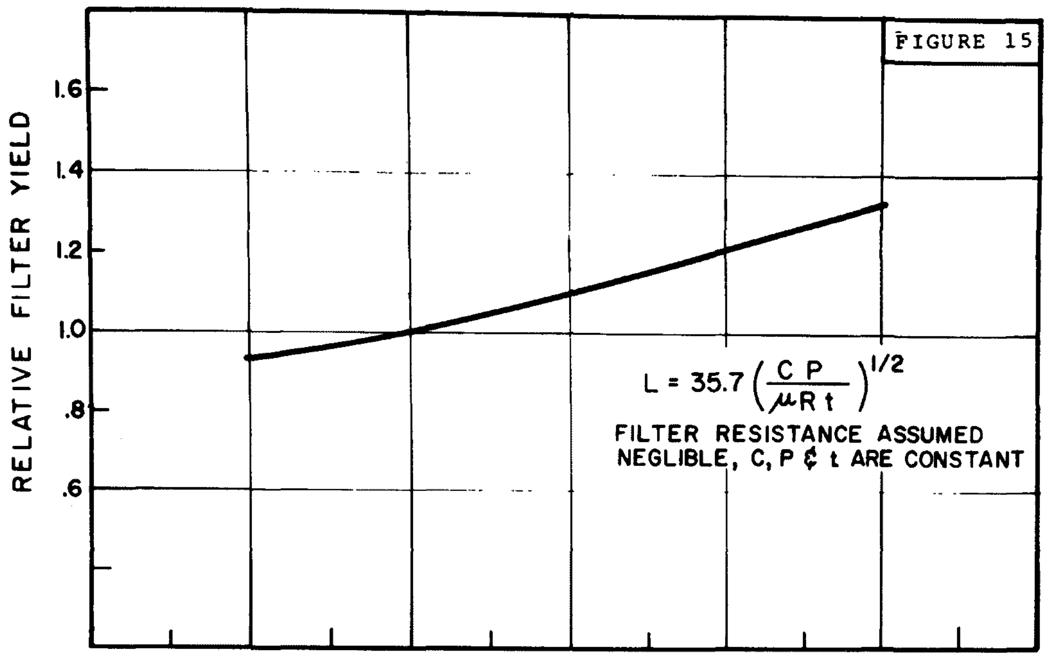
The temperature effect on pressure filtration should generally exhibit the same relationship presented for vacuum filtration.

Centrifugation

Applications of centrifuging are washing, dewatering, classification, clarification, or more usually a combination of these.

The most effective dewatering centrifuges are horizontal, cylindrical, conical, solid-bowl machines.

Centrifuges separate solids from liquids, through sedimentation and centrifugal force. Typically, sludge is fed through the center of the unit through a screw conveyor,



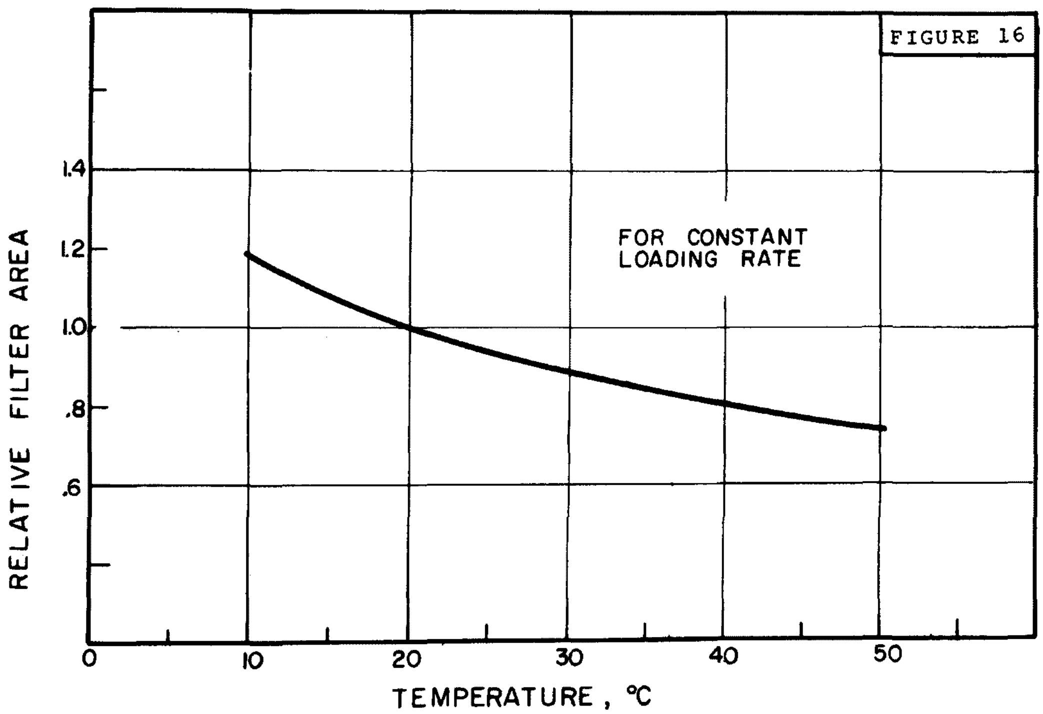


FIGURE 15 EFFECT OF TEMPERATURE ON VACUUM FILTER YIELD, CONSTANT SOLIDS LOADING.

FIGURE 16 EFFECT OF TEMPERATURE ON THE REQUIRED VACUUM FILTER SIZE TO PRODUCE A CONSTANT FILTER YIELD AT A CONSTANT SOLIDS LOADING RATE.

mounted inside a rotating conical bowl. Sludge leaving the feed tube is accelerated and is distributed to the periphery of the bowl where it is settled and compacted by centrifugal force. It is then conveyed by the screw to an inclined "beach" area where it is further dewatered and discharged. Separated liquid is discharged continuously over adjustable weirs at the opposite end of the bowl.

The factors that determine the success or failure of centrifugation are (1) cake dryness and (2) solids recovery. Guidi [34] summarized the effect of the various parameters on these two factors as follows:

Process Variable	Feed Rate	Feed Consistency	Temperature	Flocculents
To improve recovery	decrease	increase	increase	increase
To improve cake solids	increase	decrease	increase	decrease

Thus, we see that increasing temperature increases both cake dryness and solids recovery. The paramount variable though is the design of the unit itself [33].

As the temperature of the liquid carrier medium increases, the viscosity and density decrease, thus increasing the settling rate of the solids. The reduction in the moisture viscosity aids in dewatering of the solids, producing a drier cake.

The throughput capacity, Q, of a settling centrifuge at the "cutoff point" (50% of feed particles removed and 50% passed) can be described by:

$$Q = 2\,V_g \varepsilon \quad \text{.............................(\#14)}$$

with: $V_g = (\rho_p - \rho_i)d^2 g/18_\mu$ ....(#15)

and: $\varepsilon = c_\omega{}^2 r/gs$ ....(#16)

where: g = Acceleration of gravity, 981 cm/sec^2

c = Volume of liquid in bowl, cm^3

s = Effective thickness of liquid layer in which settling is occurring, cm

ρ_p = Density of particle

ρ_i = Density of liquid

d = Size of particle

μ = Absolute viscosity of liquid medium

r = Radius of curvature of path

ω = Angular velocity

Thus it can be seen that the throughput capacity for the cutoff point is inversely proportional to the viscosity and therefore directly proportional to the temperature of the liquid. An illustrative example of the theoretical effect of heat addition on centrifuge capacity is presented on Figures 17 and 18. In Figure 17, the machine and sludge operating variables are kept constant to allow determination of the theoretical effect of temperature. An increase in temperature from 20°C to 30°C is computed to approximately a 25% increase in the throughput capacity of a centrifuge. In Figure 18, the throughput rate is kept constant while the radius of curvature (r) is allowed to vary with temperature. An increase in temperature from 20°C to 30°C will result in a 20% reduction in the unit's radius of curvature. This would result in the use of a smaller diameter centrifuge for a given application.

Temperature increases may have the effect of reducing the strength of the solids and make them more difficult to convey. Ease of conveyance of a solid necessitates tests at different temperatures.

The relationships presented are necessarily theoretical since no large-scale systematic evaluation of heat effects on centrifugation has been performed.

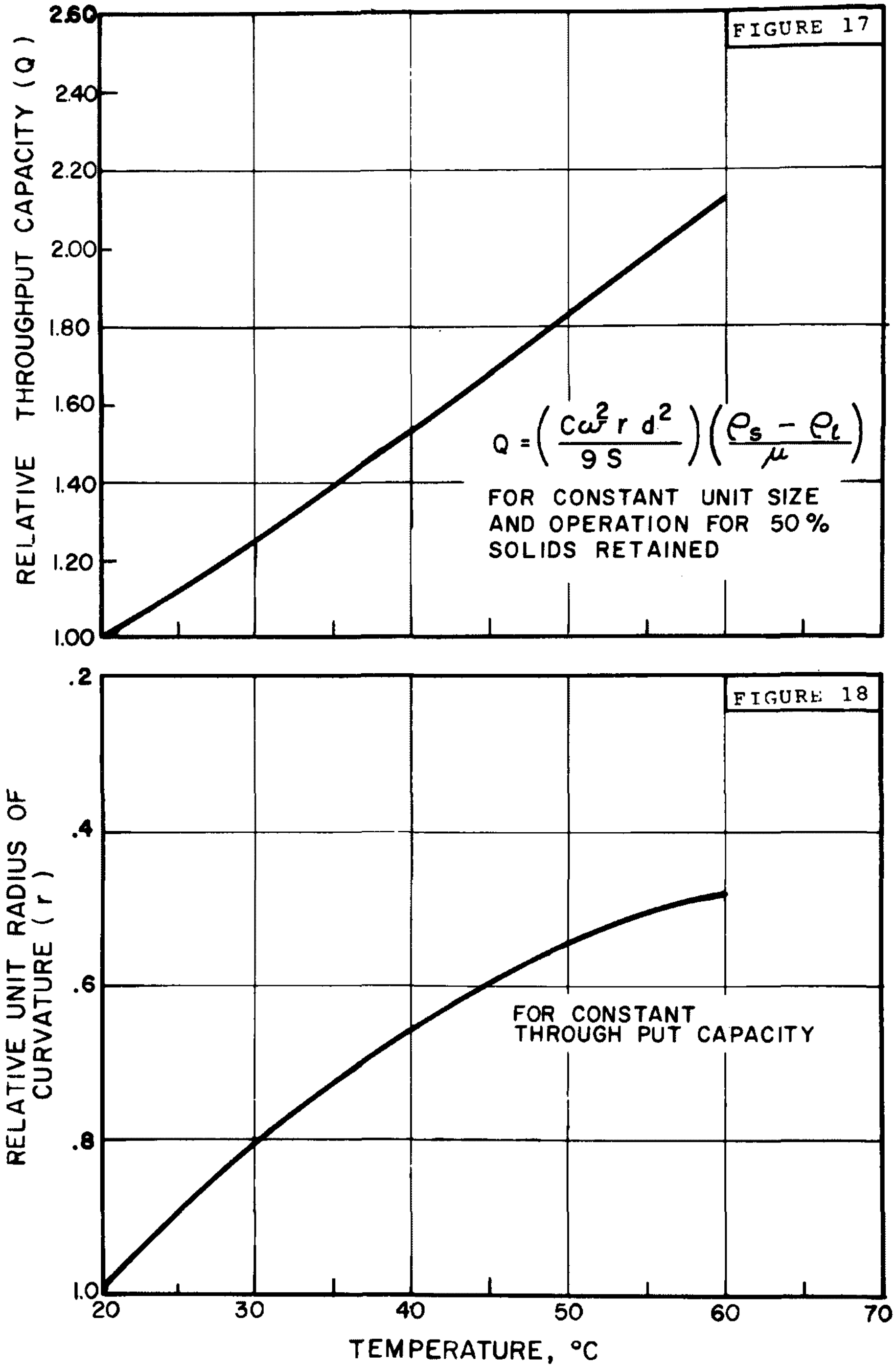


FIGURE 17 EFFECT OF TEMPERATURE ON THE CAPACITY OF A CONSTANT SIZE CENTRIFUGE.

FIGURE 18 EFFECT OF TEMPERATURE ON CENTRIFUGE SIZE TO ACHIEVE A CONSTANT THRUPUT CAPACITY.

SECTION A-IV

BIOLOGICAL PROCESSES

Biological processes are used in treatment plants to remove organic material from the wastewater and to reduce the quantity of biological sludge produced in the removal process.

Biological Waste Treatment

Biological waste treatment is used to reduce the organic materials present in a waste. It is a process wherein active bacteria are admixed with a waste. Under suitable environmental conditions, the bacteria reduce the waste to a more stable form. When the reaction proceeds in the presence of sufficient dissolved oxygen, the system is aerobic and the final decomposition products are carbon dioxide and water. Two basic phenomena occur when organic matter is removed by microorganisms: oxygen is consumed by the organisms for energy and new cell mass is synthesized. The organisms also undergo progressive auto-oxidation of their cellular mass. These reactions can be illustrated by the following general equations:

$$\text{organic matter} + O_2 + NH_3 + \text{cells} \rightarrow \text{new cells} + CO_2 + H_2O$$

and

$$\text{cells} + O_2 \rightarrow CO_2 + H_2O + NH_3$$

In the design of wastewater treatment facilities, the rate at which these reactions occur, the amount of oxygen and nutrient required, and the quantity of biological sludge produced in the reaction must be determined.

Wastes can contain suspended, colloidal, and dissolved organics. The organic matter is measured by the biochemical oxygen demand (BOD) or by the chemical oxygen demand (COD). The BOD may be defined as the amount of oxygen required by suitable organisms in the stabilization of a given quantity of organic matter. Theoretically, an infinite time is required for complete biological oxidation of organic matter, but for practical purposes, the reaction may be considered complete in twenty days. The conventional BOD test is a measure of the quantity of oxygen utilized in the first five days of oxidation, under standard conditions, and is designated as BOD_5. The quantity of oxygen required to satisfy the twenty day demand is usually referred to as ultimate BOD_u. The COD is a measure of the ultimate BOD. However, in the COD determinations, organic matter is converted to carbon dioxide and water regardless of the biological assimilability of the substances. In the analysis of data, it must be remembered that some materials which are chemically oxidized will not be biologically oxidized.

Biological waste treatment, then, essentially consists of controlling environmental factors to enable a mixed culture of microorganisms to utilize the organic matter in the waste as a food source for reproduction (synthesis) and energy (assimilation). In aerobic treatment systems, organisms are generally suspended in a liquid medium with the waste to be treated. In trickling filters, organisms are fixed to a solid medium and the waste is trickled over the medium. Dissolved oxygen is required by the culture and sufficient time is allowed for the organisms to utilize the organics as a food source.

The suspended and colloidal organic matter measured as BOD undergoes an initial reduction by adsorption to the organisms. Thereafter BOD removal is assumed to be reduced in accordance with kinetics of the first order. The removal reaction is usually expressed:

$$-\frac{dL}{dt} = KL \quad \text{..............................(\#17)}$$

which can be written in the form:

$$-\frac{dL}{dt} = K_2 S_a L \quad \text{.........................(\#18)}$$

where: $\frac{dL}{dt}$ = Rate of change of BOD with respect to time

L = BOD remaining

S_a = Quantity of microorganisms present

K_2 = BOD removal rate

The BOD removal can be expressed for more complex reactions that are retardant in nature as:

$$-\frac{dL}{dt} = K_1L_1 + K_2L_2 \ldots + K_nL_n \quad \text{..........(\#19)}$$

The reaction constant in the BOD removal equations is temperature dependent. It is possible to relate the effect of temperature on BOD removal by the following relationship:

$$K_t = K_{20}\theta^{(T-20)} \quad \text{......................(\#20)}$$

where: K_t = BOD removal coefficient at temperature T (°C)

K_{20} = BOD removal coefficient at 20°C

T = Temperature in treatment system (°C)

θ = Temperature coefficient (1.020 - 1.080)

Increases in system temperature generally increase substrate removal rates in biological treatment processes. The optimum temperature for biological reactions depends upon the type of process that is considered. For aerobic mesophilic systems (e.g., trickling filters, activated sludge and aerobic lagoons), the optimum temperature has a range between 30°C and 35°C. For aerobic thermophilic reactions (e.g., composting), the optimum temperature is approximately 52°C. For anaerobic mesophilic digestion, the optimum temperature is 37°C, while thermophilic digestors are operated at temperatures of 52°C.

In general, the system will approach some minimum BOD value rather than zero concentration due to an equilibrium between the bacteria and their liquor. The magnitude of the initial removal is a function primarily of sludge concentration, acclimatization, and waste composition. The rate of reaction is a function of temperature, nutrient level, concentration of waste, and sludge composition.

The growth of biological solids may be considered in the following steps: first, a lag period in which the culture adapts from its previous environment to the present; second, a period of maximum growth under conditions where unlimited food is available; third, a period of declining growth where food availability finally becomes a limiting condition and the sludge consumes previously stored food; and finally, an endogenous phase where, under severely limited food conditions, cells die and are, in turn, consumed so that mass population is reduced. The final sludge mass is always more than the initial, since certain non-disposable materials are generated during synthesis. The oxygen utilization rate per unit weight of sludge is low at first, but quickly reaching a maximum. As the competition for food becomes more acute, the rate decreases until an endogenous level of demand is reached. A knowledge of the parameters governing these reactions is necessary in the design of a biological system to treat any organic waste.

A general substrate removal equation can be derived from a materials balance around a completely mixed suspended growth system.

INPUT - OUTPUT - REACTION = CHANGE

$$W(t) - QC - Vf(S_a, C) = \frac{dc}{dt} V \quad \text{..........(\#21)}$$

where:

- $W(t)$ = Influent quantity (QC_o)
- C = Effluent concentration
- Q = Flow
- V = Volume of reactor
- $f(S_a, C)$ = Functional form describing biological reaction
- S_a = Active biological population measured as mixed liquor suspended solids
- C_o = Initial concentration

The reaction term can be further defined and the equation rearranged:

$$\frac{W(t)}{V} - K_2 S_a C^n = \frac{dc}{dt} + \frac{C}{t_o} \quad \text{.............(\#22)}$$

where:

- K_2 = Biological reaction rate
- n = Order of biological reaction defining dependence on substrate concentration

t_o = Detention time, V/Q

This general equation can now be used to develop the steady-state equation which assumes no change with respect to time and would be representative of a biological system operating at equilibrium under a constant organic load. The general steady-state equation reduces to:

$$\frac{(C_o - C)}{t_o} - K_2 S_a C^n = 0 \quad \text{................(\#23)}$$

In each of the foregoing equations, the term n, designating the order of the biological reaction, has been included. Kinetics describing substrate removal can generally be defined as:

$n \approx 1$ First order kinetics generally applied to BOD and COD removal. The substrate removal is directly proportional to the substrate concentration.

$n \approx 0$ Zero order kinetics generally applied to the removal of specific compounds such as linear alkyl sulfonate and phenols. The substrate removal proceeds at a fixed rate independent of the concentration.

$n \approx -1$ Retardant kinetics generally applied to bacteriostatic or inhibitory compounds such as formaldehyde. The substrate removal is inversely proportional to the concentration; removal decreases as concentration increases.

A short discussion of the specific biological processes reviewed, the pertinent literature findings, and the temperature model developed follows.

Stabilization Ponds

Waste stabilization ponds have undergone sufficient study and development to be classified as one of the major types of wastewater treatment systems. The design of a waste stabilization pond depends upon the treatment objective. A pond may be designed to receive untreated wastewaters, primary treatment plant effluents, secondary biological treatment plant effluents, or excess activated sludge [40].

Stabilization ponds rely on natural reaeration to apply oxygen to the biological populations present. Large level areas are required as well as temperate climates. The cost of stabilization ponds can be up to 50% less than equivalent activated sludge biological treatment [41].

Stabilization ponds can be generally divided into three classifications: aerobic, anaerobic, and facultative ponds. The classification depends on the organic loading and therefore dissolved oxygen content. Factors which affect a stabilization pond's efficiency include detention time, depth, organic loading, temperature, visible light energy, and the efficiency of conversion of light energy into chemical energy.

Fair et al. [6] suggest that the effluent BOD from a single stabilization pond can be described by a first-order equation as follows:

$$Y = \frac{Y_o}{1 + K_o t_d} \quad \text{..........................(\#24)}$$

where: Y_o = Influent BOD_5

Y = Effluent BOD_5

K_o = BOD removal rate constant, 1/day

t_d = Detention time, days

Gloyna [40, 41] indicates that for single pond the ratios of the reaction rates are equal to the ratios of the detention times and are a function of temperature. This relationship is:

$$\frac{k_{35}}{k_T} = \frac{t_T}{t_{35}} = \theta^{(35-T)} \quad \text{...................(\#25)}$$

where: k = Reaction rate constants for various temperatures, 1/day

t = Reaction times

T = Temperature

Laboratory data obtained by Gloyna [41] at 9°, 20°, 24°, and 35°C showed that θ ranged between 1.072 and 1.085 and K_{35} = 1.2 for a synthetic non-settleable sewage.

As pond temperature increases, the equation shows that the detention time (pond volume) requirements will decrease accordingly until the temperature reaches 35°C. The relationship between pond capacity and temperature is valid only for temperatures ranging between 3°C and 35°C. The lower limit is due to retardation of bacterial and algal activity as the temperature approaches the freezing point, while the upper limit is imposed by thermal inactivation of most types of algae.

A pond can function very well when the entire contents are not oxygenated photosynthetically. The biological degradation rate in ponds is temperature dependent. Practical design criteria necessitate careful selection of reaction rates and minimum temperatures, as well as the common considerations of light intensities, food, etc. For many domestic wastes, the following empirical relationship is suggested by Gloyna [41]:

$$V = CQL_a\ [\theta^{(35-T)}]\ f\text{-}f^1 \quad \text{..............(\#26)}$$

where: V = Pond volume (ac-ft)

Q = Influent flow (gpd)

L_a = Ultimate influent BOD (mg/l)

θ = Temperature coefficient (1.072 - 1.085)

T = Average temperature of the coldest month (°C)

C = 10.7×10^{-8} (used where temperature fluctuations are large and designs are based on a depth of 5 ft and one extra foot for solids storage)

f = Algal toxicity factor = 1 for domestic wastes

f^1 = Sulfide correction = 1 for $SO_4^=$ concentrations of less than 500 mg/l

An illustrative example showing the theoretical effect of temperature on BOD removal efficiency and required pond volume is presented on Figures 19 and 20. Figure 19 shows the decrease in BOD removal efficiency expected in a stabilization pond designed to achieve 85% removal in 7.3 days

STABILIZATION POND

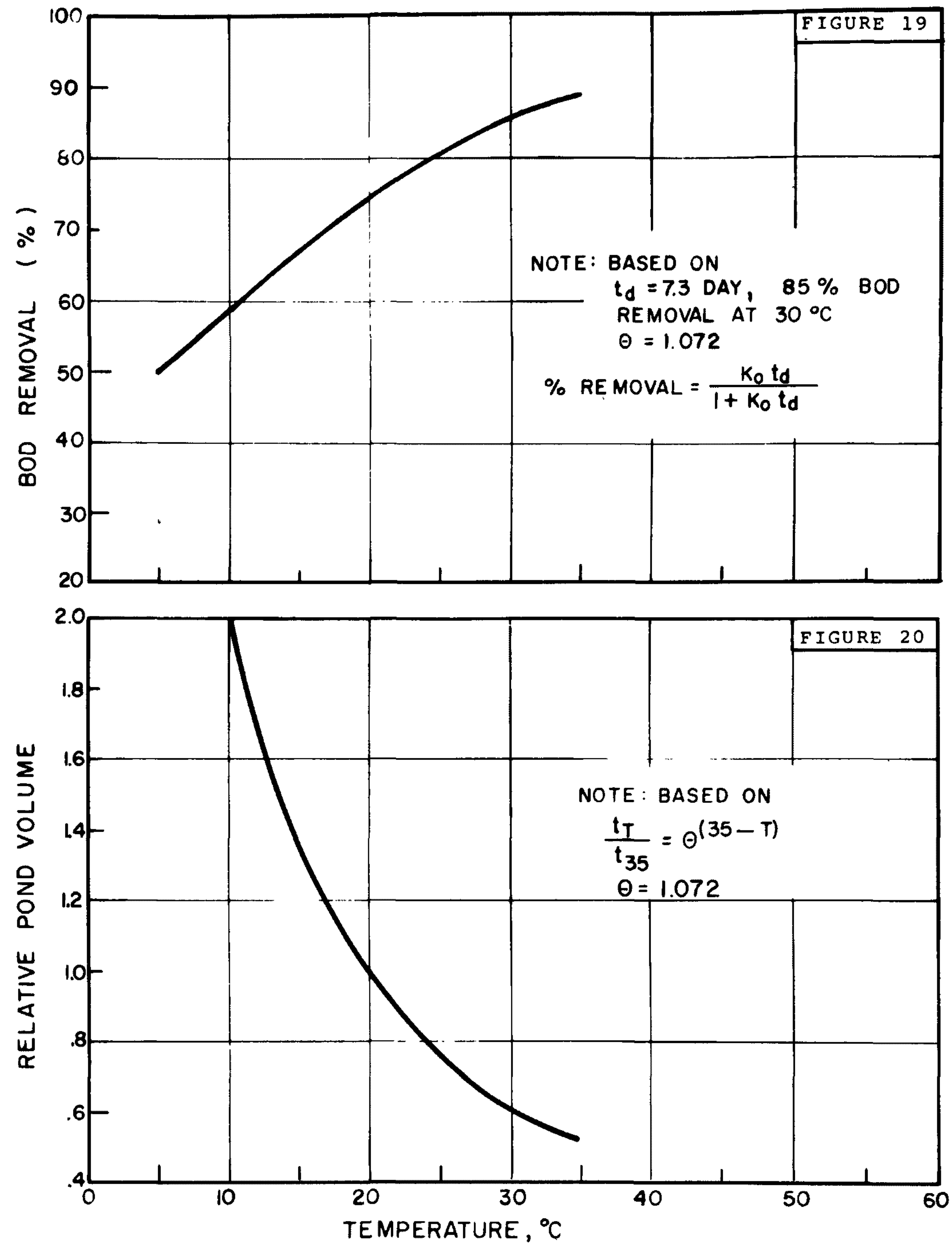


FIGURE 19 EFFECT OF TEMPERATURE ON THE BOD REMOVAL EFFICIENCY OF A CONSTANT UNIT SIZE STABILIZATION POND.

FIGURE 20 EFFECT OF TEMPERATURE ON THE REQUIRED SIZE OF A STABILIZATION POND TO PRODUCE A CONSTANT BOD REMOVAL EFFICIENCY.

at 30°C. Equations (#20) and (#24) and a θ of 1.072 are employed to develop the model shown. An increase in pond temperature from 20°C to 30°C will result in an increase in BOD removal to 85%.

Figure 20 shows the relative pond volume required to achieve a specified removal efficiency as temperature changes. Equation (#25) and a θ of 1.072 are employed to develop the model shown. All variables, except time and temperature, are kept constant in the pond evaluated. A reference temperature of 20°C is used for Figure 20. A decrease in temperature from 20°C to 10°C would require a pond of twice the size needed at 20°C to achieve an equivalent BOD removal, while by increasing the temperature from 20°C to 30°C a 40% reduction in size is possible.

Temperature variations have significant effects on stabilization ponds, as Figures 19 and 20 show. Large capital cost savings can be realized if pond temperatures can be maintained at uniformly high levels. For this reason, stabilization ponds are found mostly in the southern and southwestern areas of the United States. In colder northern climates, weather and land costs mitigate against their use.

Aerated Basins

An aerated basin (lagoon) differs from a stabilization basin in that it is usually deeper, in the order of 10 to 14 ft, and that the majority of the dissolved oxygen required for the biological processes is supplied by mechanical devices. These mechanical devices also provide the mixing in the system. The turbulence levels maintained within the basin should be sufficient to ensure a uniform dissolved oxygen concentration in the basin, but in many cases are not sufficient to maintain all suspended solids in solution. As a result, certain solids settle to the bottom of the pond, where they undergo anaerobic decomposition with the subsequent return of materials to the basin contents.

The principal drawback of the aerated lagoon system is the relatively high concentration of suspended solids leaving the system. Many recent lagoons have included baffle sections which allow the solids to settle and re-enter the mixed portion of the pond [25]. Other designs have favored clarifiers or clarification ponds following these lagoons. When internal baffling systems or external clarifier systems with sludge return are employed, aerated lagoon systems become similar to activated sludge systems.

The aerated lagoon has many desirable features. Among these are ease of operation and maintenance, equalization of the waste, and the ability to dissipate heat when desirable. The principal disadvantages of the process are the relatively large land areas required, the inability to significantly modify the process, the effluent solids loss, and the overall sensitivity of the process efficiency to changes in ambient temperature.

The rate of BOD removal may be defined by taking a materials balance around the system as described in Equation (#24). The form of the first-order equation used to describe BOD removal in aerated lagoons is [25]:

$$\frac{L_e}{L_o} = \frac{1}{1 + K_t} \quad \text{.........................(\#27)}$$

where: K_t = The product of k and the solids level

L_e = Effluent BOD concentration, mg/l

L_o = Influent BOD concentration, mg/l

This equation ignores the effect of solids sedimentation or resuspensions which are assumed to be in equilibrium.

Temperature affects the rate of biological oxidation in aerated lagoons as in other biological systems. The temperature relationship normally used to describe this phenomenon is [25]:

$$K_t = K_{20}\theta^{(T-20)} \quad \text{.....................(\#28)}$$

where: θ = Temperature coefficient

K_t & K_{20} = Biological reaction rates

Aerated lagoons have been found to be quite sensitive to temperature, with θ of 1.06 and 1.09 being reported.

An illustrative example of the effect of temperature on aerated lagoon performance has been developed based on data presented by Eckenfelder [25]. Figures 21 and 22 present the results of this example. Figure 21 presents the change in BOD removal efficiency with temperature. A base condition of 85% BOD removal in 3.5 days at 20°C with θ = 1.08

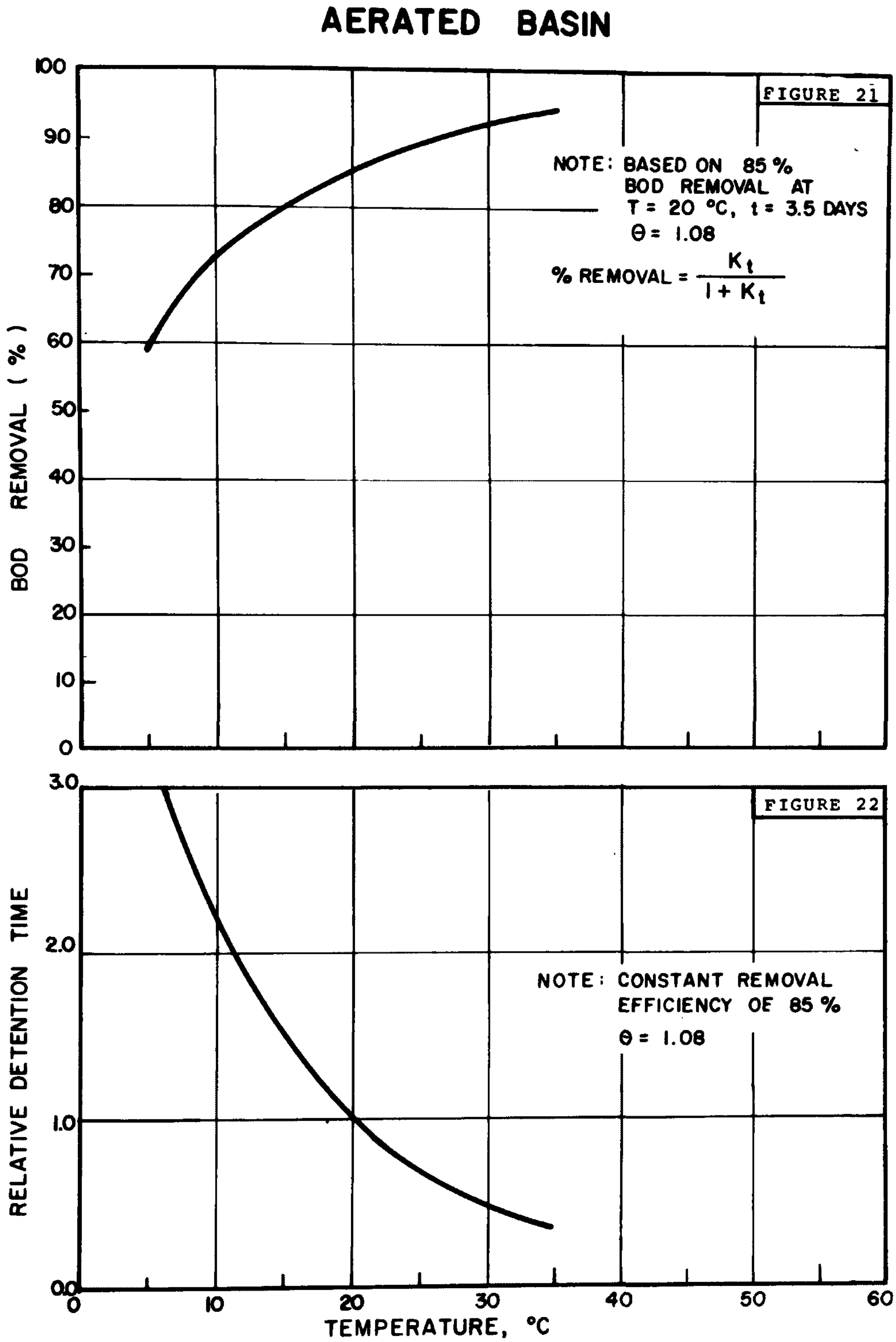


FIGURE 21 EFFECT OF TEMPERATURE ON THE BOD REMOVAL EFFICIENCY OF A CONSTANT SIZE AERATED BASIN.

FIGURE 22 EFFECT OF TEMPERATURE ON THE REQUIRED SIZE OF AN AERATED BASIN TO ACHIEVE A CONSTANT BOD REMOVAL EFFICIENCY.

is used for the example. An increase in temperature from 20°C to 30°C will result in an increase in BOD removal efficiency to 92%. Figure 22 presents the relative detention time required to achieve 85% BOD removal as the temperature changes. An increase in temperature from 20°C to 30°C will require a basin approximately 40% smaller than that required to achieve 85% BOD removal at 20°C.

The temperature of a waste will change during treatment in a lagoon system. The temperature which will be maintained in the lagoon system will depend on the heat balance between the influent wastewater and the ambient air temperature. Heat is lost through evaporation, convection, and radiation, and is gained by solar radiation. The total heat loss from the lagoon may be defined by the following relationship [42]:

$$H = H_e + H_c + H_r - H_s \quad \ldots\ldots\ldots\ldots\ldots\ldots (\#29)$$

where: H = Net heat loss

H_e = Heat loss by evaporation

H_c = Heat loss by convection

H_r = Heat loss by radiation

H_s = Heat gain by solar radiation

The heat loss due to evaporation, H_e, is expressed by the relationship [42]:

$$H_e = 0.00722\ H_v C\ (1 - 0.1W)(V_w - V_a) \quad \ldots (\#30)$$

where: H_v = Latent heat of vaporization, Btu/hr-SF

C = Constant characteristic of the lagoon

W = Mean wind velocity, mph

V_w = Vapor pressure at the liquid surface

V_a = Vapor pressure at the atmosphere

The heat loss by convection, H_c, is computed from the relationship:

$$H_C = (0.8 + 0.32W/2)\ (T_w - T_a) \quad ....(\#31)$$

where: T_w = Lagoon temperature (°F)

T_a = Air temperature (°F)

The heat loss by radiation, H_r, can be expressed by the relationship:

$$H_r = 1.0\ (T_w - T_a) \quad ....(\#32)$$

The net heat gain by solar radiation has not been defined at this time from available data and is neglected in these calculations.

Activated Sludge

The activated sludge process may be defined as a system in which flocculated biological growths are continuously circulated and contacted with organic wastes in the presence of oxygen. The oxygen and mixing in the system are supplied by diffused air, mechanical aeration, or a combination of the two. The process involves an aeration step followed by a solids-liquid separation step. The separated solids are returned for admixture with the waste in the aeration phase. Many modifications of the process are employed to obtain the desired degree of treatment. They range from high-rate processes with aeration detention times of 2 hrs to low-rate systems with aeration times of 24 hrs or more, depending on waste strength.

There are several models which relate BOD removal efficiency with temperature. However, these models are applicable only in the mesophilic temperature range. The basic form of the equation describing the BOD removal efficiency in a completely mixed, activated sludge process has been presented by Eckenfelder [5, 25]:

$$E_T = \frac{K_T S_a t L_e}{L_a} \quad ....(\#33)$$

where: E_T = BOD removal efficiency at temperature T°C

K_T = BOD reaction coefficient at temperature T°C

S_a = Mass of biological volatile solids

t = Detention time in aeration tank

L_e = Effluent BOD concentration

L_a = Influent BOD concentration

The van't Hoff-Arrhenius relationship for the effect of temperature (in a certain range) on purification rates can be used for analysis, provided that the experimental determination of the θ value for the sewage at different temperature ranges is obtained. This relationship is expressed in the following equation [5, 25]:

$$K_T = K_{20}\ \theta^{(T-20)} \qquad \text{(\#34)}$$

where: K_{20} = BOD reaction coefficient at 20°C

θ = Temperature coefficient

T = Temperature in °C

The optimum temperatures for the biological oxidation of various types of waste have been reported in the 30°C to 52°C range. A summary of these investigations is presented on Figure 23. For each investigation reported, the temperature range and optimum temperature are indicated.

The elevated temperature used in the activated sludge process depends upon the optimum value for biological reaction obtained experimentally for the specific wastewater involved. Based on the information presented on Figure 23, the optimum temperature for mesophilic biological systems appears to be in the 30°C to 37°C range. A temperature coefficient (θ) for activated sludge systems has been reported to average 1.02 at temperatures up to 30°C [25].

There are limited data published for biological oxidation of wastes in the thermophilic range. A thermophilic biological population does exist which is capable of BOD removals at temperatures up to 65°C [25]. However, this high temperature operation requires very stringent control over system temperature since these cultures are very temperature-sensitive.

Many factors affect the complex process of organic removal by microorganisms and it is not surprising to see different

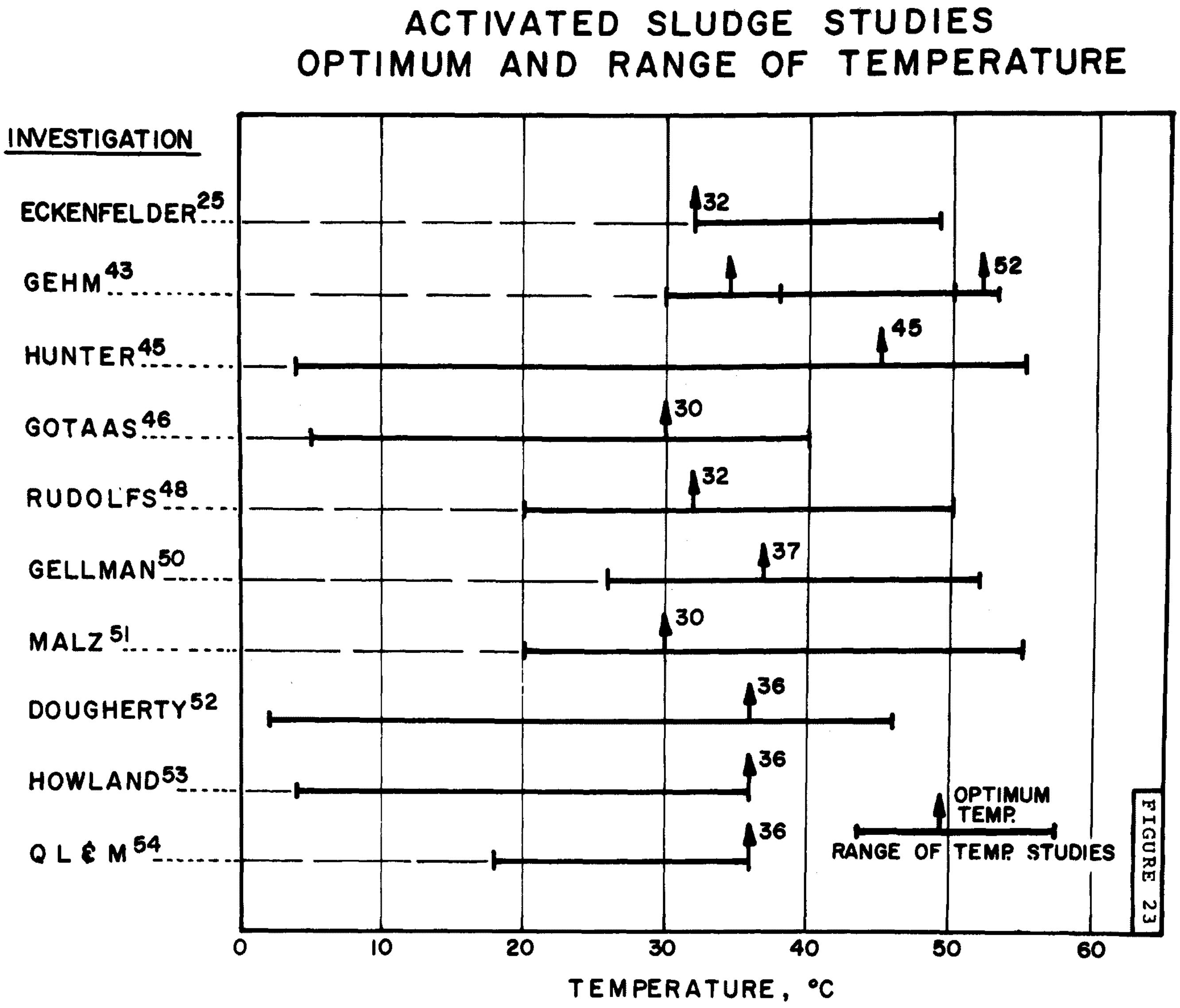


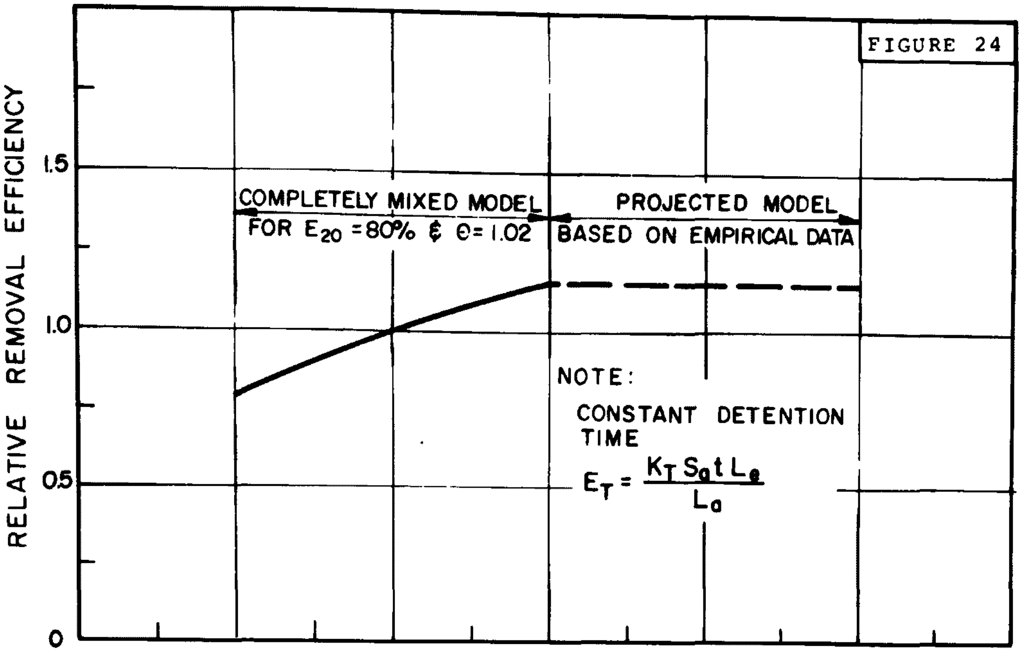
FIGURE 23 OPTIMUM TEMPERATURES FOR THE ACTIVATED SLUDGE PROCESS REPORTED IN THE LITERATURE.

investigators reported different results. One study [48] shows an efficiency decrease with rising temperature in the 30°C to 40°C range, followed by increasing efficiency until reaching an optimum value between 50°C and 55°C. This phenomenon can be explained by the fact that after the optimum mesophilic temperature range, the mesophilic bacterial population dies off, followed by growth of a thermophilic bacterial population. Other data [25, 46] show a decrease of efficiency after the mesophilic optimum is reached, which continues through the thermophilic range. A pilot study [49] of paper mill wastewater treatment reports similar efficiency at optimum mesophilic and thermophilic temperatures. It has been concluded that the activated sludge process can be operated successfully at feed temperatures as high as 52°C.

As an illustration of the effect of temperature on the activated sludge process, a model is proposed based on Equation (#33). Model performance is described by Equation (#33) up to a temperature of 30°C based on a base condition of 80% BOD removal at 20°C and a $\theta = 1.02$. At temperatures above 30°C and below 50°C, model performance remains constant at the BOD removal level achieved at 30°C. This model is proposed in light of the range of optimum performances, shown on Figure 23, most of which were from laboratory-scale evaluations. Experimentation in the 30°C plus range (up to the thermophilic range) is required to develop a specific temperature model.

Figures 24 and 25 present the anticipated effect of temperature on the relative removal efficiency and detention time. Figure 24 indicates that an increase in temperature from 20°C to 30°C will result in an additional 20% increase in removal of the remaining substrate. In the example used, this would correspond to an 84% BOD removal at 30°C in an activated sludge system designed for 80% BOD removal at 20°C. Figure 25 presents the relative detention times required to achieve a constant BOD removal or the relative % substrate remaining. Also presented on Figure 25 are Rudolf's data [48] for the temperature range of 30°C to 50°C. These data substantiate the proposed model (constant removal efficiency) in that temperature range. In this model, a temperature increase from 20°C to 30°C would result in a system with 30% less detention time to achieve the same BOD removal.

In order to keep BOD removal at optimum rate, sufficient amounts of dissolved oxygen are needed in the activated



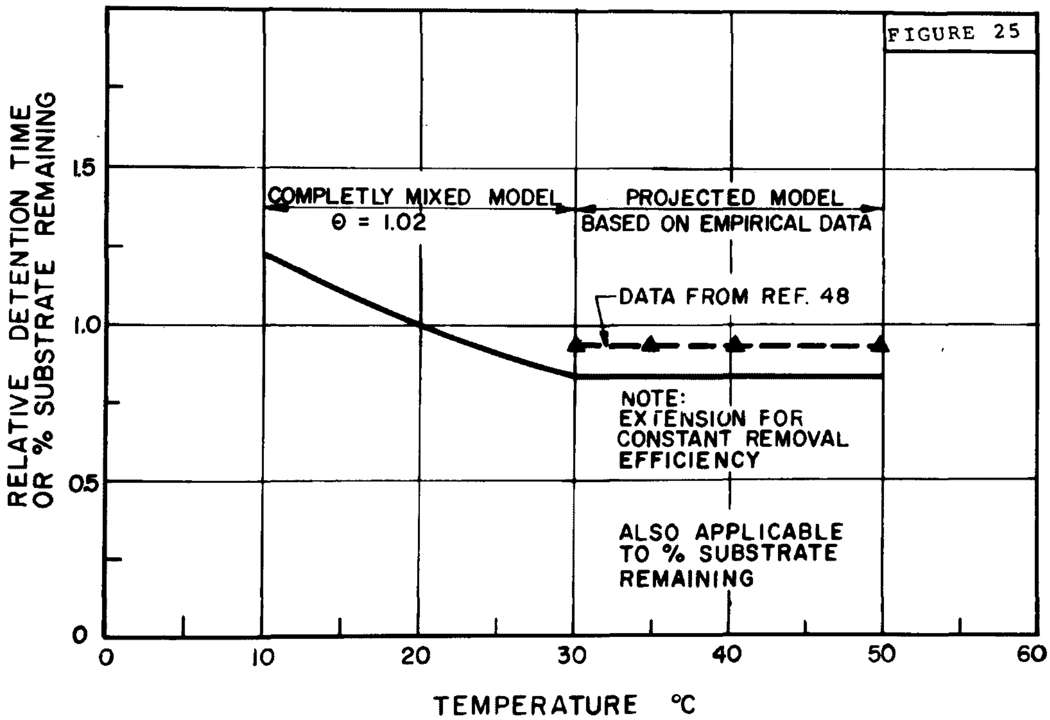


FIGURE 24 EFFECT OF TEMPERATURE ON BOD REMOVAL EFFICIENCY OF A CONSTANT SIZE COMPLETELY MIXED ACTIVATED SLUDGE UNIT.

FIGURE 25 EFFECT OF TEMPERATURE ON THE REQUIRED ACTIVATED SLUDGE UNIT SIZE TO PRODUCE A CONSTANT BOD REMOVAL EFFICIENCY.

sludge process. Dissolved oxygen saturation varies with temperature. At elevated wastewater temperatures, the oxygen utilization rate of microorganisms increases and the saturation of dissolved oxygen decreases. The driving force (difference between saturation and actual concentration) for oxygen transfer from air to wastewater decreases with increasing temperature. A check of the theoretical oxygen requirements of an activated sludge system operating over a temperature range of 10°C to 50°C has been made. The difference in aeration requirements over the temperature range of 10°C to 50°C is within 10%. This would be within the design safety factor normally used when selecting oxygenation equipment.

Trickling Filters

A trickling filter is a packed bed of medium (e.g., plastic, stone) covered with microbial slime. As the wastes pass over the slime layer, the organic material present in the waste is reacted upon by microbial action.

Similar to the activated sludge process, the van't Hoff-Arrhenius relationship can be applied for predicting the temperature effect on trickling filter performance. Howland determined a $\theta = 1.035$ from his experimental data.

An illustrative example of the effect of temperature on trickling filter performance is presented on Figure 26. Howland's filter equation and temperature coefficient are used in this example. A base condition of 75% BOD removal at 20°C and a constant detention time are also used. Figure 26 indicates that a temperature increase from 20°C to 30°C will result in a 15% increase in relative removal efficiency from 75% BOD removal at 20°C to 86% BOD removal at 30°C.

Rotating Disks

Biological fixed-film rotating disk (BFFRD) is a treatment process which involves the use of biological films attached to the rotating disk. The experimental results reveal that the BFFRD is an efficient treatment process for organic removal, ammonia removal, and resisting organic shock loadings [60]. The limiting factors in substrate removal are the dissolved oxygen content, diffusion, or substrate concentration. The process variables include influent loading, flow rate, detention time, temperature, number of stages of disk, surface area, submerged depth, speed, and direction.

TRICKLING FILTER

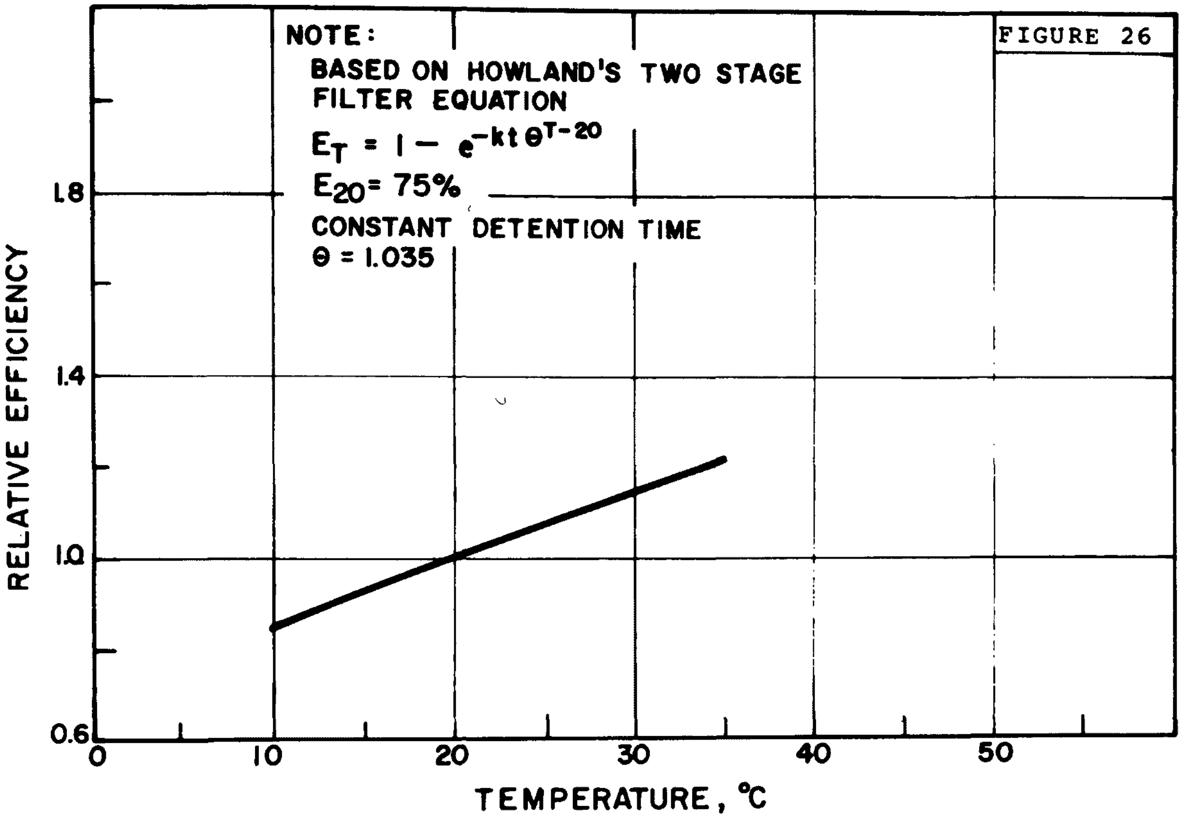


FIGURE 26 EFFECT OF TEMPERATURE ON THE BOD REMOVAL EFFICIENCY OF TRICKLING FILTERS, UNIT SIZE CONSTANT.

Mathematical models describing the BOD removal efficiency are reported [60]. Temperature is one of the variables included in these models. Experimental data indicate that BOD removals increase with increasing temperature. However, experimental data are limited to the mesophilic range of 20°C to 30°C. A temperature coefficient for BOD removal rate of $\theta = 1.025$ has been determined for a single-stage BFFRD.

Anaerobic Digestion

Anaerobic digestion is commonly used to stabilize the waste sludge generated during municipal waste treatment. Principally, the sludges are comprised of primary sludge and excess sludge from either the activated sludge or the trickling filter operations.

Digestion produces a sludge more amenable to dewatering, disposal, lagooning, dilution, and other disposal methods [19]. Digestion occurs in a mixed culture of microorganisms where particular species are most active in different stages. The decomposition is accompanied by gasification, liquefaction, stabilization, colloidal structure breakdown, and the release of moisture. Since the digestion process is not complete, byproducts of intermediate metabolism include organic acids, ammonia, methane, hydrogen sulfide, carbon dioxide, and carbonates [61]. Volatile solids reduction of about 70% is commonly achieved by anaerobic digestion [18] in the mesophilic range with approximately 24 days' detention time.

One of the more important factors controlling the rate and completeness of digestion is temperature [62]. Numerous studies [19,63,64,65,66,67] describe the effect of temperature on these design parameters and models. There are two distinct ranges for operation of anaerobic digestors. Low-range temperatures are called mesophilic and cover the range of about 88°F to 103°F (31°C to 40°C). The thermophilic temperatures range from approximately 116°F to 132°F (46°C to 55°C). There is some disagreement concerning the exact maximum and minimum temperature of these ranges, as the following table shows:

Thermophilic*			Mesophilic*				
Opt.	Max.	Min.	Opt.	Max.	Min.	Reference	Date
		108				68	1934
128			98			65	1937
122			86			66	1948
	133	122		108	90	69	1953
				100		70	1956
128	130	115	98	100	80	19	1959
	135	122		104	84	5	1961

* *All temperatures in °F.*

For the purpose of this discussion, 128°F (53°C) and 98°F (37°C) will be used for the optimum temperature for thermophilic and mesophilic operation, respectively. In each range a corresponding bacteria population is responsible for digestion.

Thermophilic operation offers various advantages over mesohilic operation. Thermophilic temperatures result in significantly smaller digestion periods [19]. The capacity of the digestor is directly proportional to the time of digestion, thus indicating proportional decrease in required capacity in the thermophilic ranges. Reduction in required digestor capacity of 33% to 50% is suggested by A.S.C.E. [19] for thermophilic operations, independent of concentration and storage. A capacity temperature relationship based on the work of Fair and Moore [65] is presented on Figure 27. Loadings for digestors run at thermophilic ranges were found to be 1.6 times the maximum loading in the mesophilic range for the same density of sludge [71].

Digestors run at thermophilic temperatures (128°F) are reported to be stable and resistant to upset [72]. However, because of operating difficulties few plants utilize the process. Doubling of solids loading of the digestor was found to have little effect at thermophilic temperatures, except to reduce gas production and volatile solids reduction a small amount [72].

Gas production at thermophilic temperatures is approximately 10% higher than at mesophilic temperatures despite smaller digestion capacities [62,64,68,71]. Golueke [62] evaluated gas production in laboratory-scale digestors. Results of his studies are presented on Figure 28. Golueke also found that the gas components are essentially the

ANAEROBIC DIGESTION

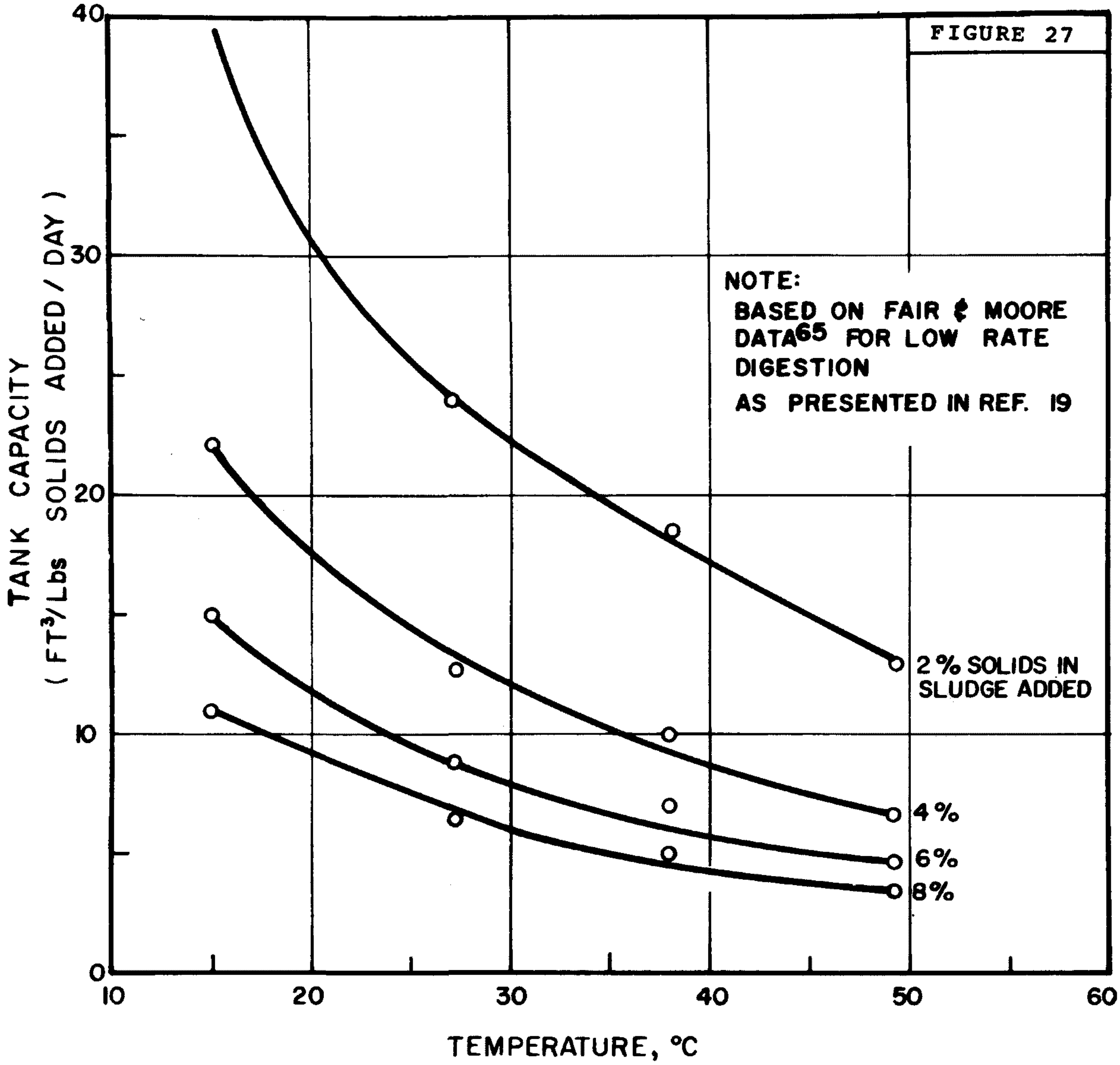


FIGURE 27 EFFECT OF TEMPERATURE ON ANAEROBIC DIGESTOR SIZE REQUIRED TO PRODUCE A CONSTANT VOLATILE SOLIDS DESTRUCTION AT VARIOUS SOLIDS CONCENTRATION.

ANAEROBIC DIGESTION

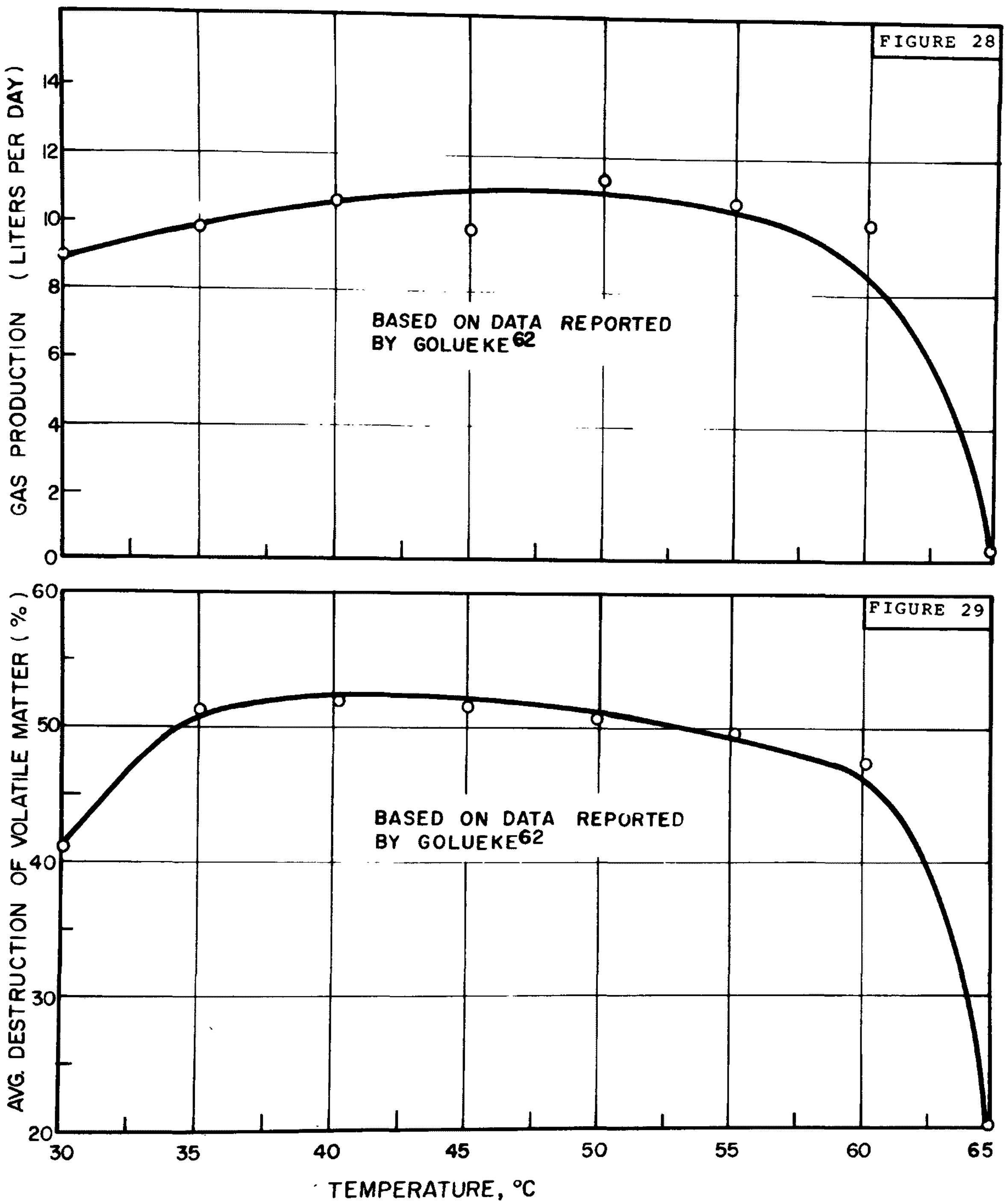


FIGURE 28 EFFECT OF TEMPERATURE ON GAS PRODUCTION IN ANAEROBIC DIGESTORS OF CONSTANT SIZE.

FIGURE 29 EFFECT OF TEMPERATURE ON VOLATILE SOLIDS DESTRUCTION EFFICIENCY IN CONSTANT SIZE ANAEROBIC DIGESTORS.

same, while some investigators found an increase in methane from thermophilic digestors. Maly [67] found a greater organic nitrogen conversion to ammonia at thermophilic ranges, while Garber [72] found less total nitrogen in the digested sludge.

The efficiency of the digestor in destruction of volatile matter decreases by approximately 5% at thermophilic temperatures [62]. Figure 29 presents the data developed by Golueke in laboratory-scale studies of the effect of digestor temperature on the destruction of volatile matter.

Improved sludge handling characteristics for thermophilic sludges were found by Garber [72] and supported by Golueke [62]. These changes in characteristics were found to be: (1) the average particle size was larger; (2) the proteinaceous material was more completely digested; and (3) the sludge had less total nitrogen. Digested sludge concentrations increased from 3.64% to 4.85% and average vacuum filter yields increased from 1.7 to 6.3 lbs/SF/hr [72]. Chemical dosage (ferric chloride) was also reduced from 6.5% to 3.4% in Garber's study.

Popova [73] found that most pathogenic microbes (except viruses) are destroyed at 50°C and that all viable eggs of helminths were gone at a digestion temperature of 51°C. Rawn [74] found no problems with odor in his studies when live steam was used to heat digestors to the thermophilic range.

Digestor supernatants can impose a high BOD and solids load on other treatment plant processes and the effluent receiving water. Fischer [71] found that the supernatant from thermophilic digestors contained more solids, being high in colloidal and non-settleable solids which were difficult to remove even at high coagulant doses of lime and ferric chloride. Two-stage digestion, a mesophilic first stage followed by a thermophilic second stage, resulted in a supernatant of better quality. Golueke [62] states that Fischer's findings of poor supernatant quality could have been due to inadequately adapted digestors which produced poor digestion.

Aerobic Digestion

Aerobic digestion can be described as a process where microorganisms obtain energy by auto-digestion of the cell protoplasm and the biologically degradable organic matter in the sludge cells is oxidized to carbon dioxide, water, and ammonia [75]. Aerobic digestion produces a biologically stable sludge suitable for a variety of further dewatering and disposal operations [18].

Volatile solids reduction is dependent upon detention time with a sharp increase in volatile solids reduction as the detention time is extended to about 12 days [76].

Lawton [76] and Drier [77] found that temperature has an appreciable effect at short detention times and that the effects of temperature decreased at longer detention times or lower loading rates. Figure 30 presents the relationships developed by Drier [77] in his study of temperature effects on aerobic digestion. These results are based on bench-scale studies of mixed primary and waste-activated sludge. Continuous flow reactors, with feed sludge at 3.2% solids concentrations and approximately 70% to 80% volatile solids, were used by Drier.

Drier [77] also found that pH and alkalinity in the aerobic digestor rose with increasing temperature. Lawton [76] found that long detention times were required to produce a digested sludge with good settling and dewatering characteristics. Loehr [78] and Woodley [79] found that thermophilic (52°C) aerobic digestors were less efficient than those operating in the mesophilic (35°C) range.

AEROBIC DIGESTION

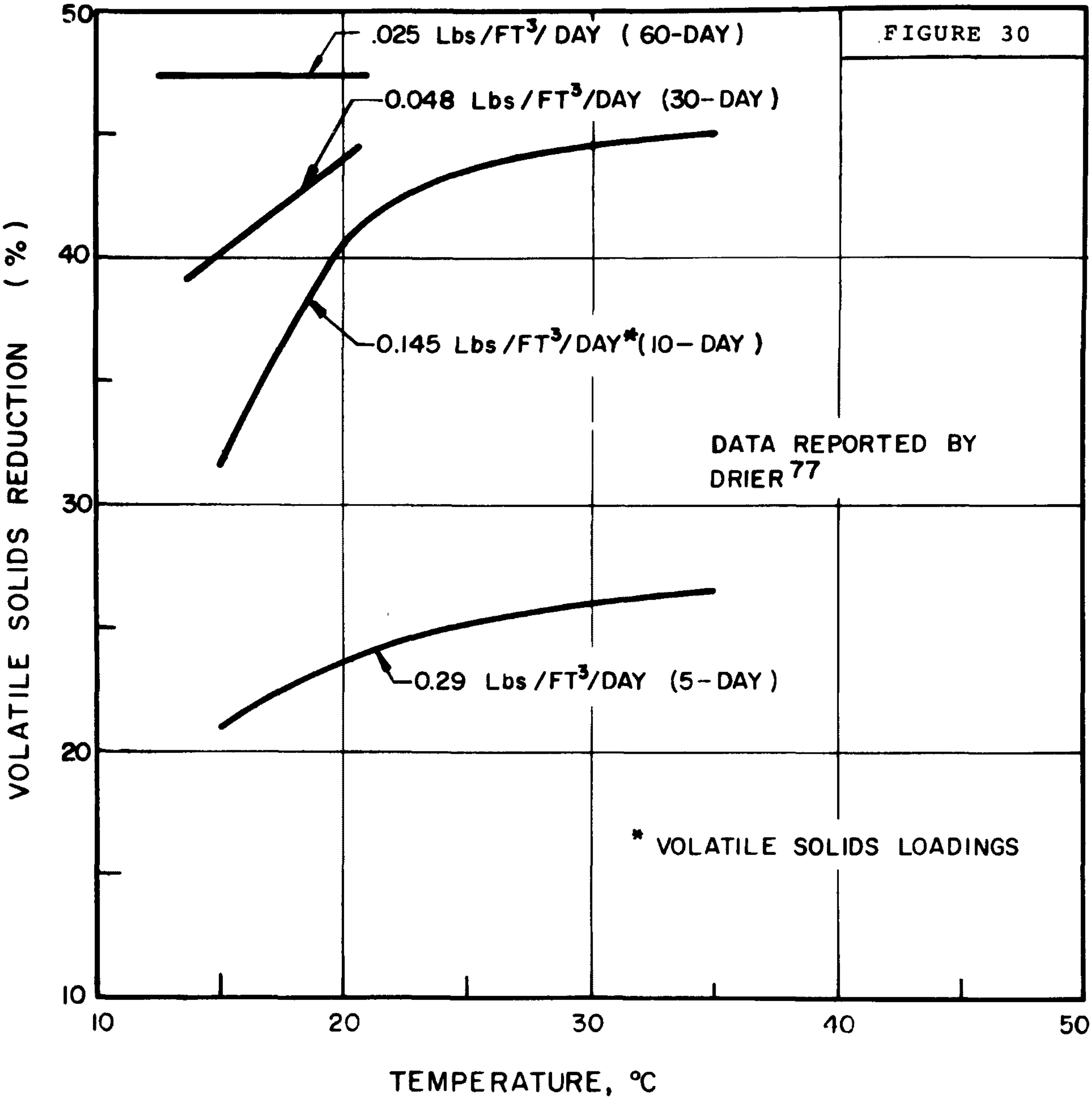


FIGURE 30 EFFECT OF TEMPERATURE ON VOLATILE SOLIDS DESTRUCTION EFFICIENCY IN CONSTANT SIZE AEROBIC DIGESTORS.

SECTION A-V

DISINFECTION

Chlorination

Chlorination has long been considered to have the greatest practical potential of all disinfection systems for freeing sewage of pathogens [80]. Chemical disinfection theoretically proceeds in two steps: (1) penetration of the cell wall, and (2) reaction with the cell enzymes [6].

The rate of disinfection is generally considered to be determined by the rate of diffusion of the disinfectant through the cell wall or the rate of reaction with an enzyme. The van't Hoff-Arrhenius relationship can be used to describe temperature effects [6].

A convenient form for this relationship is:

$$\log \frac{t_1}{t_2} = \frac{E\ (T_2 - T_1)}{4.56\ T_1 T_2} \quad \ldots\ldots\ldots\ldots\ldots\ldots (\#35)$$

where: T_1 & T_2 = Two absolute temperatures (°K) for which the rates are to be compared

t_1 & t_2 = Times required for equal percentages of kill at fixed concentrations of disinfectant

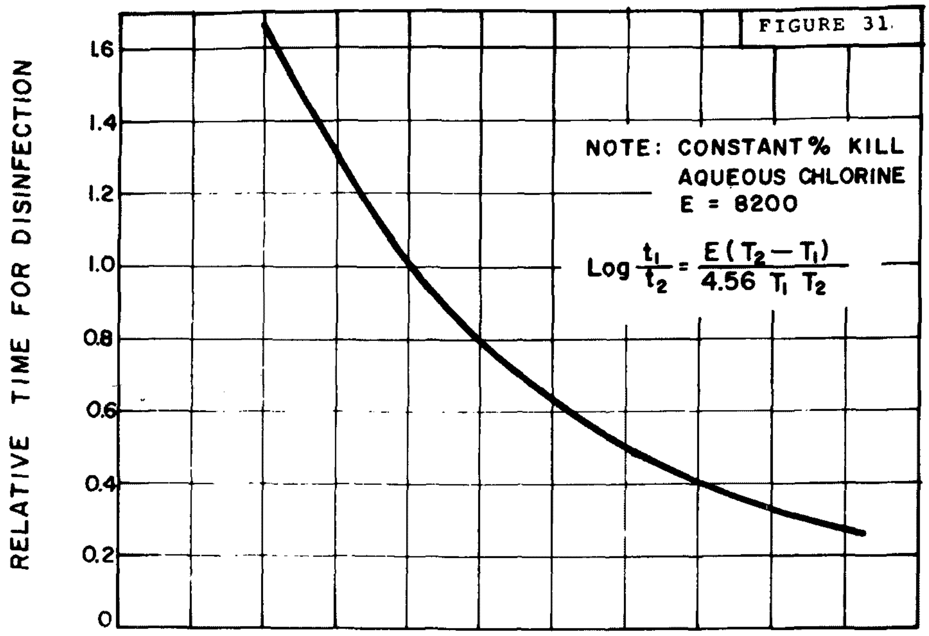
E = Activation energy (calories) and is a constant characteristic of the reaction (for aqueous chlorine, E = 8,200 @ pH 7.0)

For $T_2 - T_1 = 10$, the ratio t_1/t_2 (called Q_{10}) is approximately related to E at normal water temperature as follows [6]:

$$\log Q_{10} = \log (t_1/t_2) = E/39{,}000 \quad \ldots\ldots (\#36)$$

For aqueous chlorine at a wastewater pH of 7.0, E = 8,200 and Q_{10} = 1.65. An illustrative example of the theoretical temperature effect on the contact time required to achieve a constant percent kill using aqueous chlorine is presented on Figure 31. A temperature increase from 20°C to 30°C

CHLORINATION



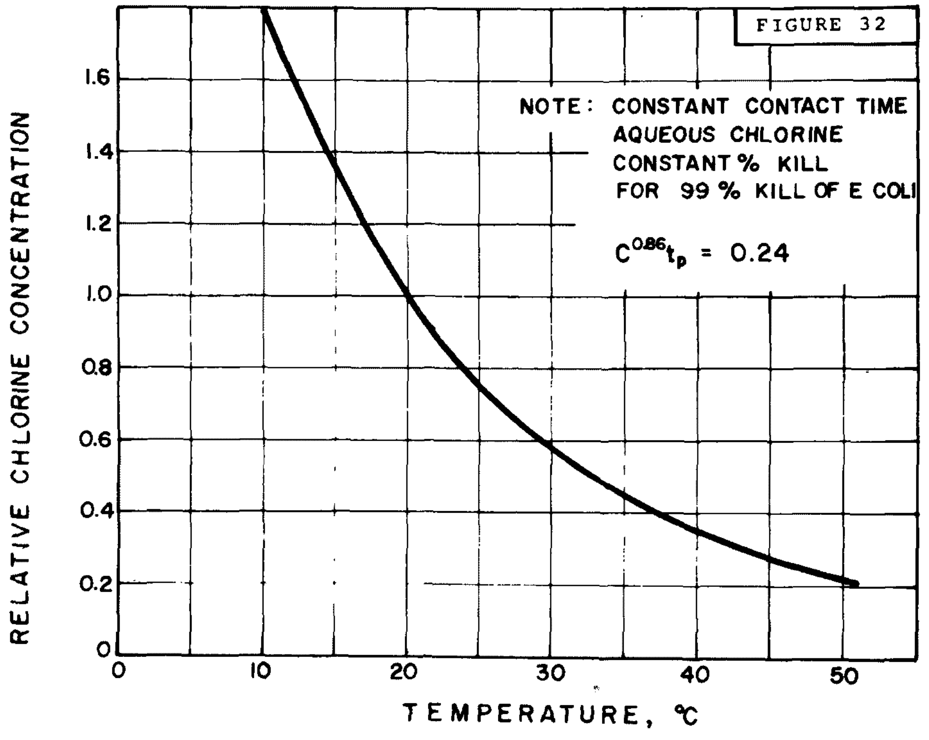


FIGURE 31 EFFECT OF TEMPERATURE ON THE TIME REQUIRED TO PRODUCE A CONSTANT PERCENT KILL WITH A CONSTANT CHLORINE RESIDUAL.

FIGURE 32 EFFECT OF TEMPERATURE ON THE CHLORINE REQUIRED TO PRODUCE A CONSTANT PERCENT KILL.

will result in approximately a 35% reduction in the required contact time for disinfection.

A relationship generally used to describe the observed disinfection efficiency as a function of disinfectant concentration is [6]:

$$C^n t_p = \text{Constant} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (\#37)$$

where: C = Concentration of disinfectant

t_p = Time required to effect a constant % kill

n = A coefficient of dilution or a measure of the order of the reaction

For a 99% kill of E. coli with aqueous chlorine as HOCl, Berg [6] has found the following relationship:

$$C^{0.86} t = 0.24 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (\#38)$$

Using this equation and the information on Figure 31, a relationship between required chlorine dosage and temperature can be developed. Figure 32 presents the relationship between chlorine dosage and temperature for a 99% kill of E. coli with aqueous chlorine in a constant contact time. An increase in temperature from 20°C to 30°C will result in a 42% reduction in the amount of chlorine required for disinfection.

SECTION A-VI

ADVANCED TREATMENT PROCESSES

The unit operations discussed in this section have been classified as advanced treatment processes. Generally, these processes are not commonly employed in "conventional" secondary wastewater treatment. Therefore, they are classified as advanced which is arbitrary in some cases.

Many of these processes could be included solely or as part of a system to provide tertiary treatment after conventional secondary biological treatment. For most of the processes reviewed, insufficient information is available to allow development of a model describing temperature effects on the process.

Ultra-High Rate Filtration

The ultra-high rate (UHR) filtration process involves the filtration of wastewaters through a multi-media bed up to 20 ft in depth at application rates of from 12 to 30 gpm/SF [93]. Chemical coagulation prior to filtration may be employed to enhance removal efficiencies and decrease the head loss through the filter.

UHR has been applied as a tertiary treatment step to achieve higher degrees of removal of suspended solids, BOD, and phosphates. Experimental tests have been reported using alum and polymer additions to achieve suspended solids, BOD, and total phosphate removal efficiencies of 99%, 97%, and 98%, respectively [93]. The average effluent concentrations of suspended solids, BOD, and total phosphates were reported as two, four, and less than one mg/l, respectively. Further studies on the removal of other contaminants are required to establish the total effectiveness of the UHR filtration process.

The temperature effect on the UHR filtration process is expected to exhibit a similar relationship to that presented for the gravity filtration process.

Organic Carbon Removal

Activated carbon is used in wastewater treatment to remove soluble organic compounds from solution. The wastewater is normally pumped through packed bed or expanded bed activated carbon columns to remove residual refractory organic compounds from biological secondary treatment effluents. Recently activated carbon adsorption has also been used for the treatment of effluents from physical-chemical processes.

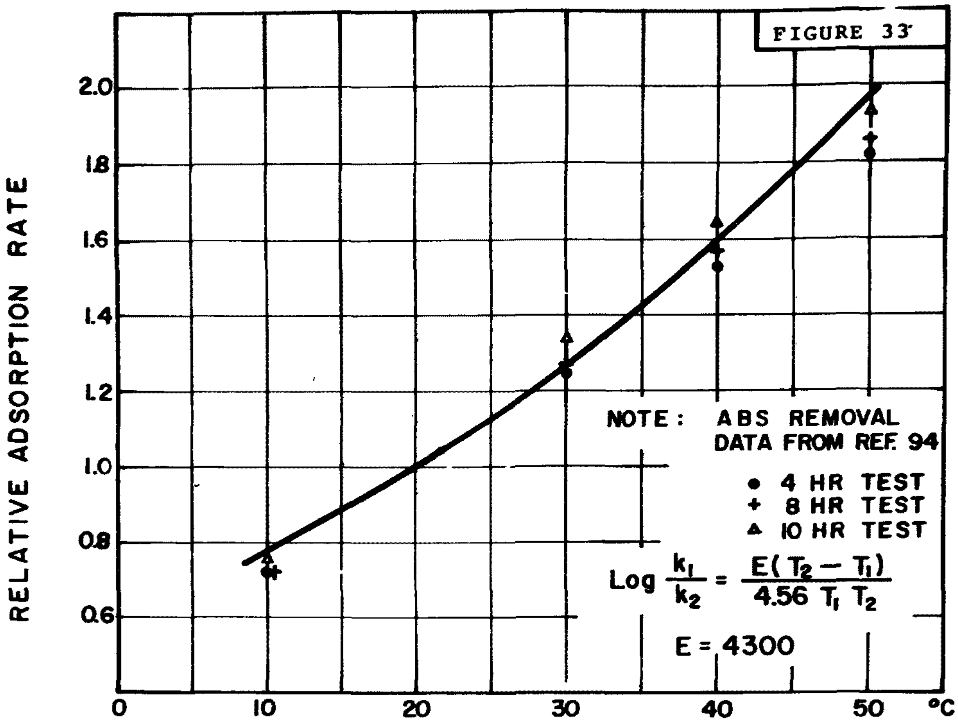
Carbon adsorption is a surface phenomenon where molecules are adsorbed due to the attraction between surface charges within the carbon pores and the adsorbate. The overall rate of adsorption is generally considered to be limited by the rate of diffusion of molecules into the carbon pores. Therefore, Equation (#35), the van't Hoff-Arrhenius equation, can be used to describe the effect of temperature on the overall rate of adsorption.

For liquid adsorption systems, there is no precise method for selecting and predicting the performance of activated carbon types founded on their basic properties or those of the adsorbing material. Design data must be obtained from pilot plant testing for the particular application under consideration.

As an illustration of the effect of temperature on carbon adsorption, the work of Morris and Weber [94] is used to develop the relationship presented in Figures 33 and 34. A series of carbon adsorption test runs were performed at various temperatures on water containing alkyl benzene sulfonate (ABS). Figure 33 presents the change in the relative adsorption rate as temperature increases. The van't Hoff-Arrhenius equation is presented along with data reported by Morris and Weber [94]. A temperature increase from 20°C to 30°C will result in a 25% increase in the rate of ABS adsorption on activated carbon.

Figure 34 presents the change in relative adsorption capacity of activated carbon for ABS as temperature increases. The Langmuir equation and data presented by Morris and Weber [94] for ABS are used to develop the relationship shown. A temperature increase from 20°C to 30°C will result in an 8% reduction in the relative adsorption capacity of the activated carbon. A specific set of design conditions would be needed to demonstrate the total value of increasing the temperature in an activated carbon system. Where high volumes of water are to be treated, an increase in adsorption rate and therefore reduction in liquid detention time would be desirable.

ACTIVATED CARBON ADSORPTION



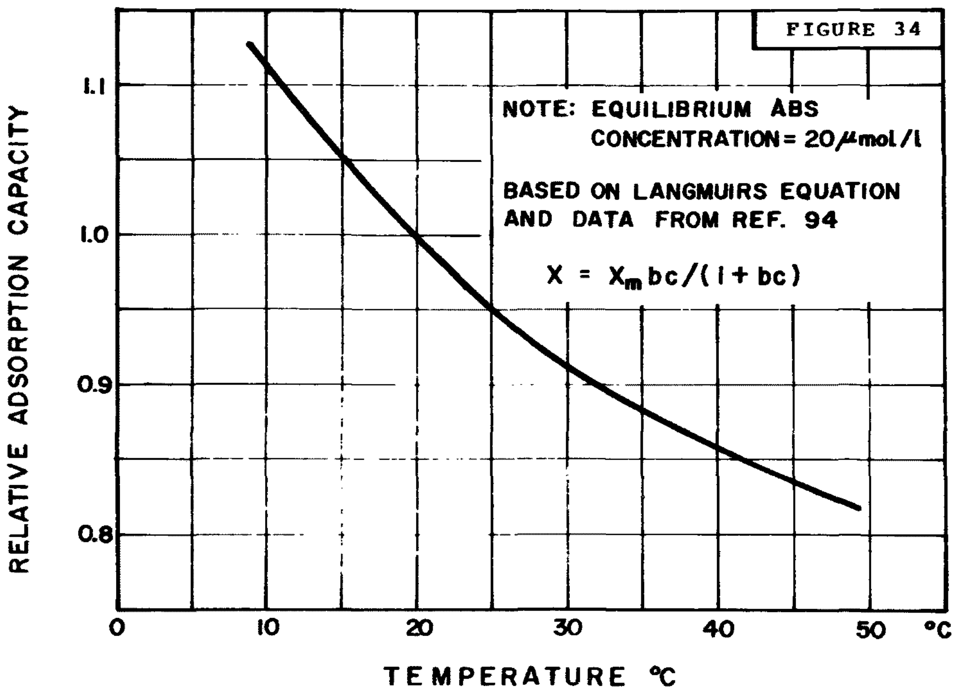


FIGURE 33 EFFECT OF TEMPERATURE ON THE ADSORPTION EFFICIENCY OF ACTIVATED CARBON WITH A CONSTANT ABS APPLICATION.

FIGURE 34 EFFECT OF TEMPERATURE ON ADSORPTION CAPACITY OF ACTIVATED CARBON WITH A CONSTANT ABS APPLICATION.

Nitrogen Removal

Physical, chemical, and biological processes have been used to remove nitrogen compounds from wastewaters. Present water quality criteria may require 90% nitrogen removal and/or a specified nitrogen content in the effluent from wastewater treatment plants.

The principal forms of nitrogen present in sewage are ammonia and organic nitrogen. Organic nitrogen is generally present in suspended form. Therefore, those processes which effect suspended solids removal also effect a removal of organic nitrogen. Ammonia nitrogen can be removed by air stripping or by conversion to another more oxidized form of nitrogen (nitrite or nitrate). The oxidized nitrogen forms can be reduced to nitrogen gas by bacterial action.

In addition to sedimentation, air stripping, and the biological processes, other more sophisticated nitrogen removal processes may be employed including electrodialysis, reverse osmosis, ion exchange, and distillation. Nitrogen removal processes discussed herein are:

- Activated sludge
- Anaerobic columnar filters
- Ammonia stripping

Activated Sludge

Nitrogen removal via air activated sludge process is a two-stage phenomenon: nitrification (the oxidation of nitrogen forms to nitrite and nitrate) and denitrification (the reduction of nitrite and nitrate to nitrogen gas).

Nitrification occurs in two steps. The first is oxidation of ammonia nitrogen to nitrite and the second is oxidation of nitrite to nitrate [95].

The pH range for the oxidation of ammonia to nitrite is 7.5 to 9.0 and the range for oxidation of nitrite to nitrate 8.0 to 9.0. The growth rate of Nitrosomonas has been described by the following [95, 96]:

$$K_m = 0.18\, e^{0.12(T-15)} \quad \ldots\ldots\ldots\ldots\ldots (\#39)$$

where: K_m = Growth rate

T = Temperature (°C)

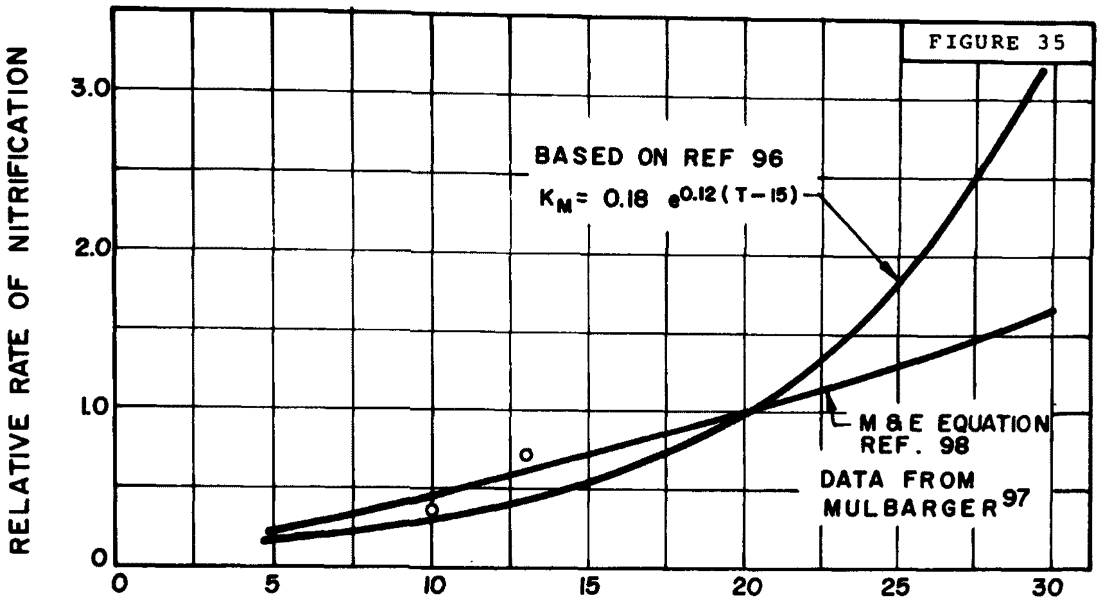
Experimental evaluations of the rate of nitrification [96, 97,98,99] show variations away from this formula. Figure 35 presents the data of Mulbarger [97] and a relationship developed by Metcalf and Eddy [98]. The relative rate of nitrification is presented as a function of temperature. Above 25°C, Mulbarger's data show a wide divergence from Equation (#39). Figure 36 presents the data of Metcalf and Eddy relating the rate of ammonia nitrogen nitrification to temperature at various pH values. Based on the information presented on these two figures, it is apparent that nitrification is severely limited by low wastewater temperatures.

Figure 37 presents the permissible nitrification tank loadings based on the work of Sawyer at Marlboro, Massachusetts [98]. The effect of lower wastewater temperatures on nitrification rate can be offset somewhat by increasing the mixed liquor volatile suspended solids, as shown on Figure 37. Where wastewater temperatures drop below 65°F (18°C), such as in northern climates, the required sludge age may be excessively high for operation of a single carbonaceous removal and nitrification stage, necessitating sequential staging [98]. However, with heat enrichment, a single sludge system for carbonaceous removal and nitrification may be sufficient and thereby provide a definite cost savings.

Nitrites and nitrates are biologically reduced to nitrogen gas in the denitrification step by a wide variety of common facultative bacteria. Denitrification can occur through the endogenous respiration of the biomass or through the addition of an organic carbon source to increase the denitrification rate and reduce the required residence time. Various organic compounds have been used as a carbon source, but methanol has been found to be the least expensive.

No workable relationship for describing the effect of temperature on the rate of denitrification in activated sludge has been developed, though data show increase of rates with higher temperature [97, 100]. Figure 38 presents Mulbarger's [97] data for denitrification rates as they vary with temperature. A rise in temperature from 10°C to

NITRIFICATION
(ACTIVATED SLUDGE)



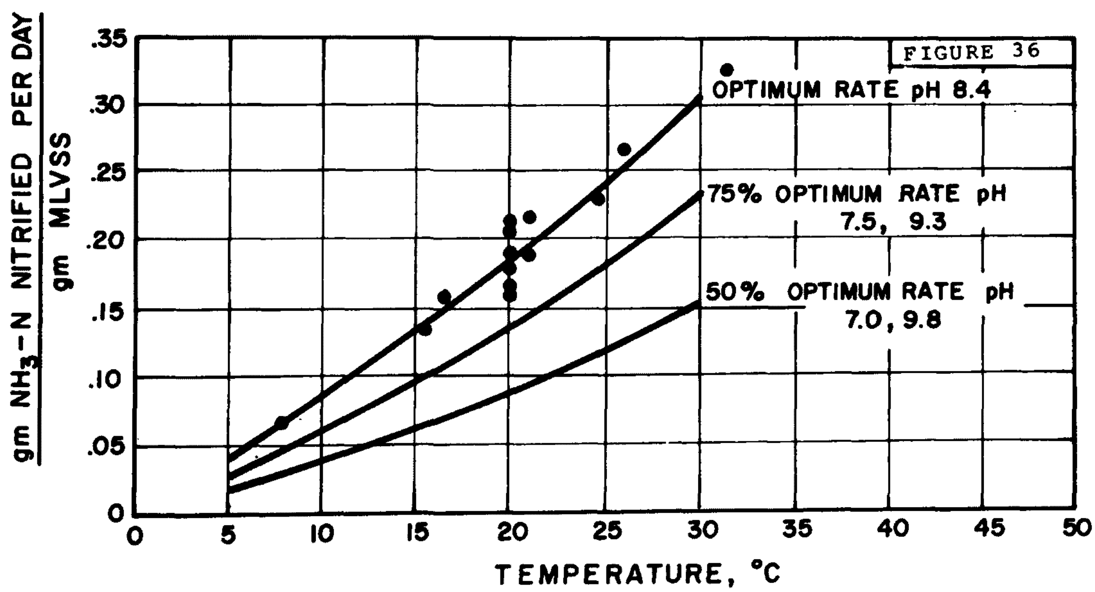


FIGURE 35 & 36 EFFECT OF TEMPERATURE ON THE RATE OF NITRIFICATION IN THE ACTIVATED SLUDGE PROCESS.

PERMISSIBLE NITRIFICATION TANK LOADINGS

DATA FROM REF. 98

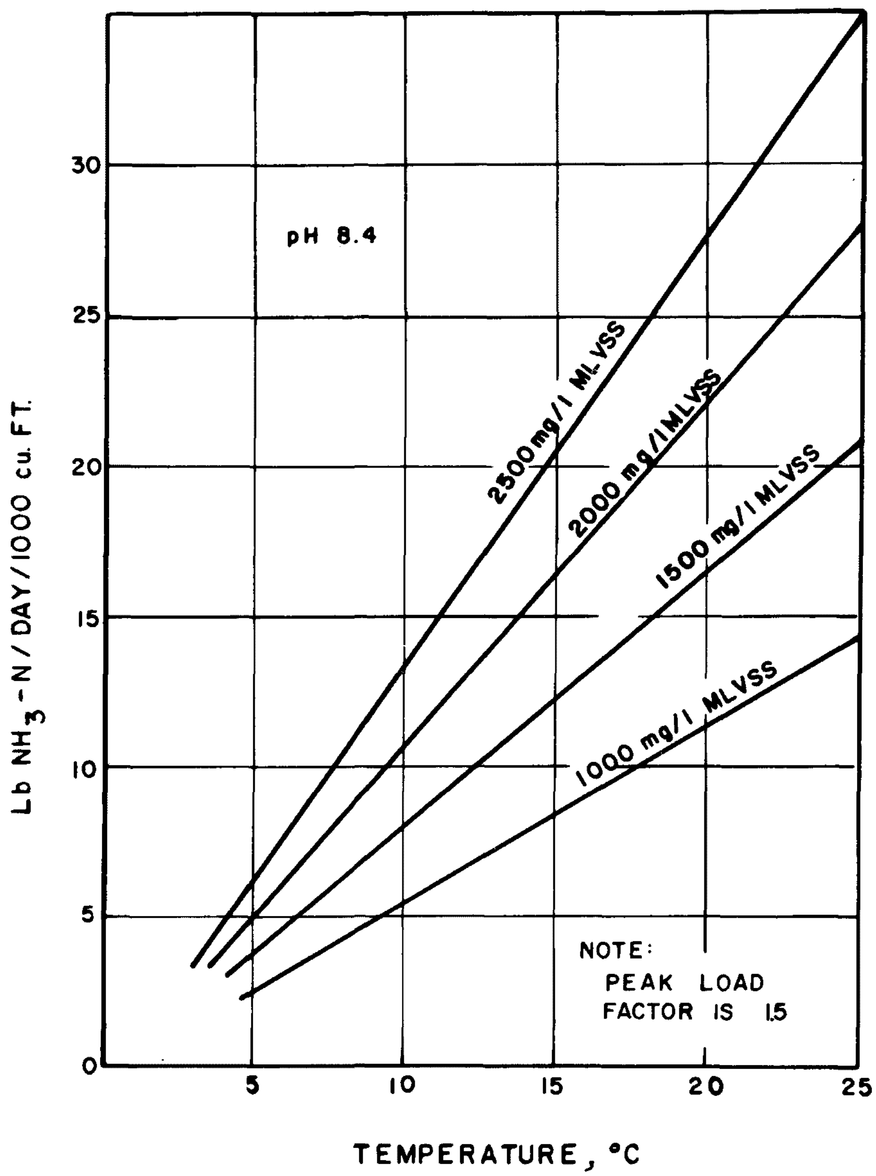


FIGURE 37 EFFECT OF TEMPERATURE ON NITRIFICATION LOADING AT OPTIMUM RATE.

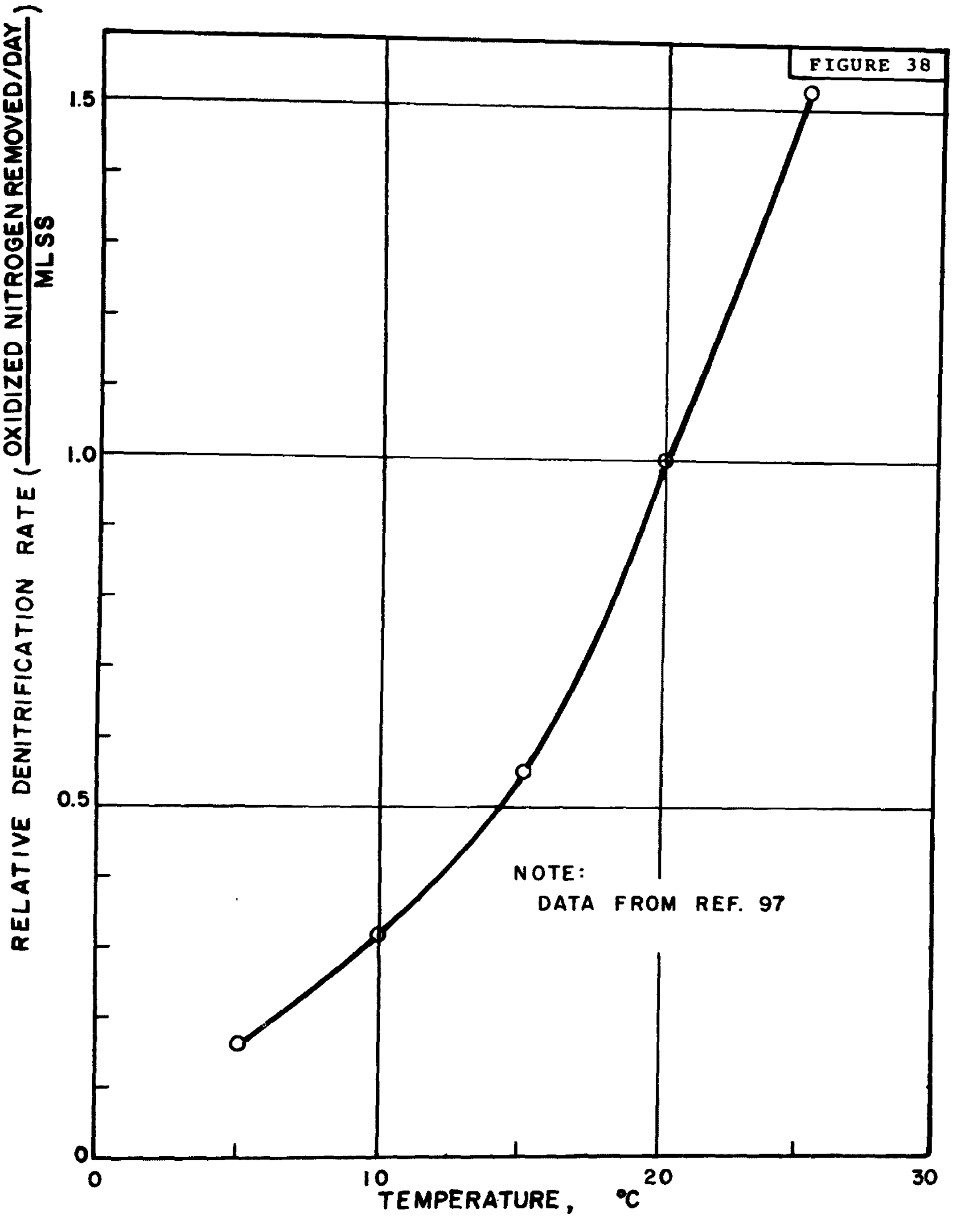


FIGURE 38 EFFECT OF TEMPERATURE ON DENITRIFICATION RATE IN THE ACTIVATED SLUDGE PROCESS.

20°C resulted in a 300% increase in the rate of denitrification. Figure 39 presents the relationships developed by Sawyer [98] in the Marlboro, Massachusetts studies.

Anaerobic Columnar Filters

Denitrification of fully nitrified waste streams may be achieved by the anaerobic filtration process. Using methanol as the hydrogen acceptor, denitrifying organisms within the adhering filter surface slime growth will reduce nitrate nitrogen to molecular nitrogen gas.

Recent experiments have shown the denitrification capabilities of the anaerobic filtration process [101]. However, mathematical models of the anaerobic filtration process have not as yet been confirmed through sufficient application.

Temperature effects were found to be small. Increasing temperature has been shown [102] to increase slightly the denitrification rate and decrease slightly the required methanol doses. Figure 40 presents the effects of temperature on nitrate nitrogen removal for various methanol doses.

The columnar contacting denitrifying system provides construction cost savings over suspended growth reactors and has the added advantage of suspended solids removal and ease of process control [103].

Ammonia Stripping

Air stripping is used to remove ammonia nitrogen from wastewaters. Ammonia stripping efficiency increased considerably with increased temperature [84]. A mathematical model has been developed which relates ammonia stripping rates with temperature [104]. Ammonia removals experienced at Lake Tahoe increased in excess of 10% for a temperature rise from 15°C to 20°C. Figures 41 and 42 present the Lake Tahoe data on ammonia stripping [84]. Figure 41 presents the relative ammonia removal rate as a function of temperature. Figure 42 presents the required stripping tower depth, to achieve a stated percent ammonia removal, as a function of temperature.

PERMISSIBLE DENITRIFICATION TANK LOADINGS

DATA FROM REF. 98

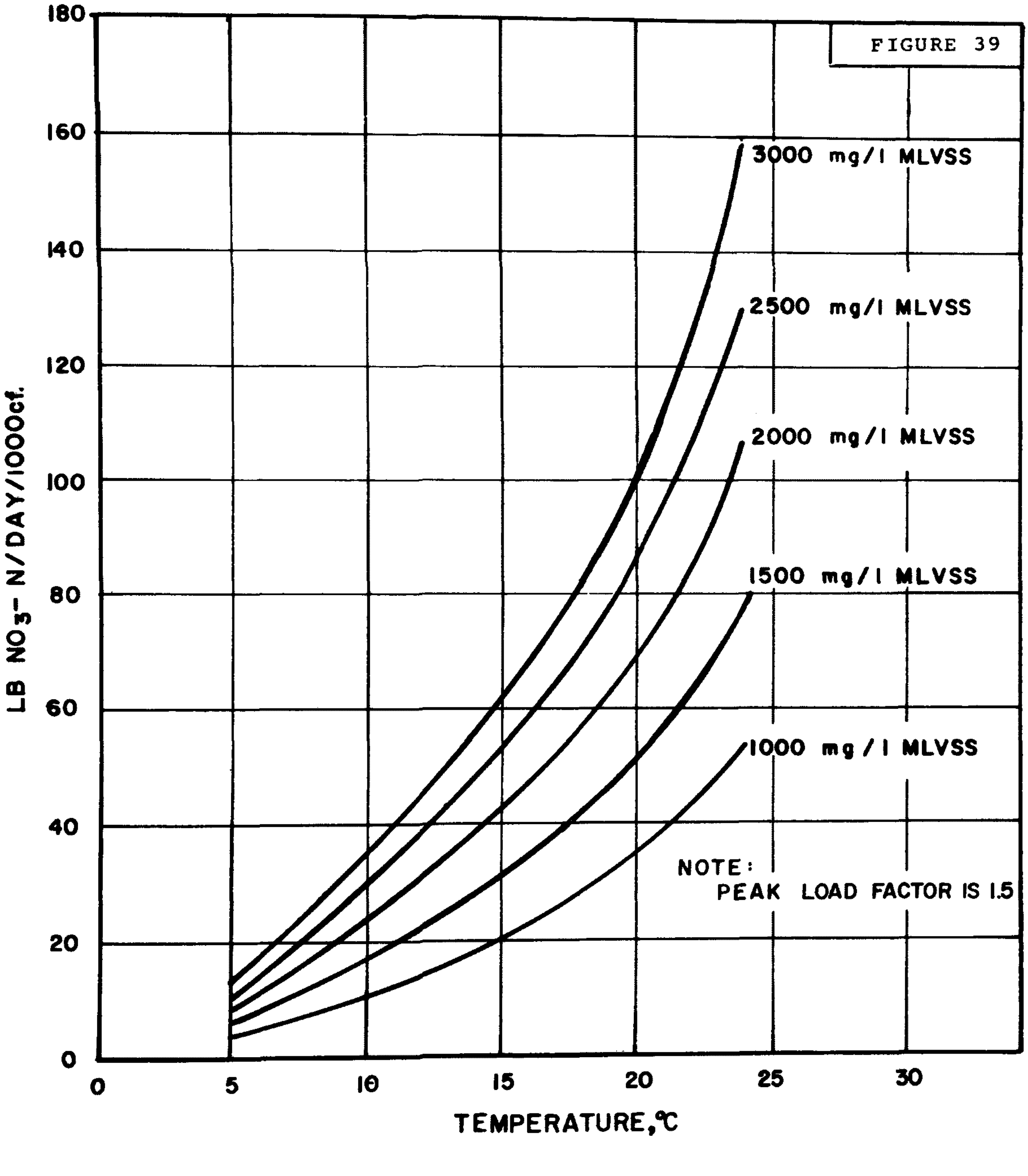


FIGURE 39 EFFECT OF TEMPERATURE ON DENITRIFICATION LOADING RATE.

ANAEROBIC COLUMNAR DENITRIFICATION

BASED ON DATA REPORTED
IN REFERENCE 102

NOTE: PARTS METHANOL / PART NO_3 —N ON
WEIGHT BASIS GIVEN IN PARENTHESIS

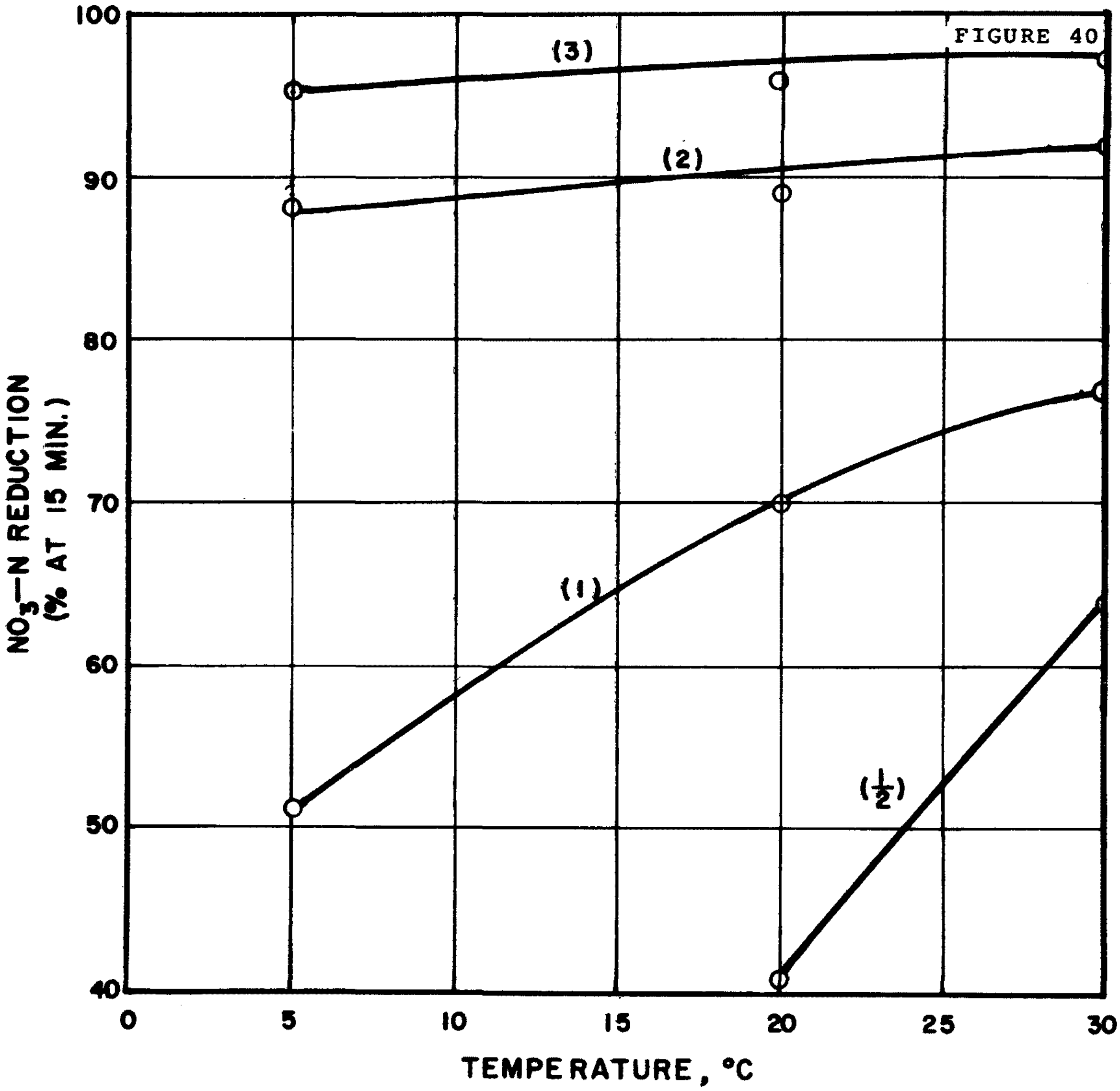


FIGURE 40 EFFECT OF TEMPERATURE ON DENITRIFICATION EFFICIENCY FOR CONSTANT SIZE DENITRIFYING COLUMNS.

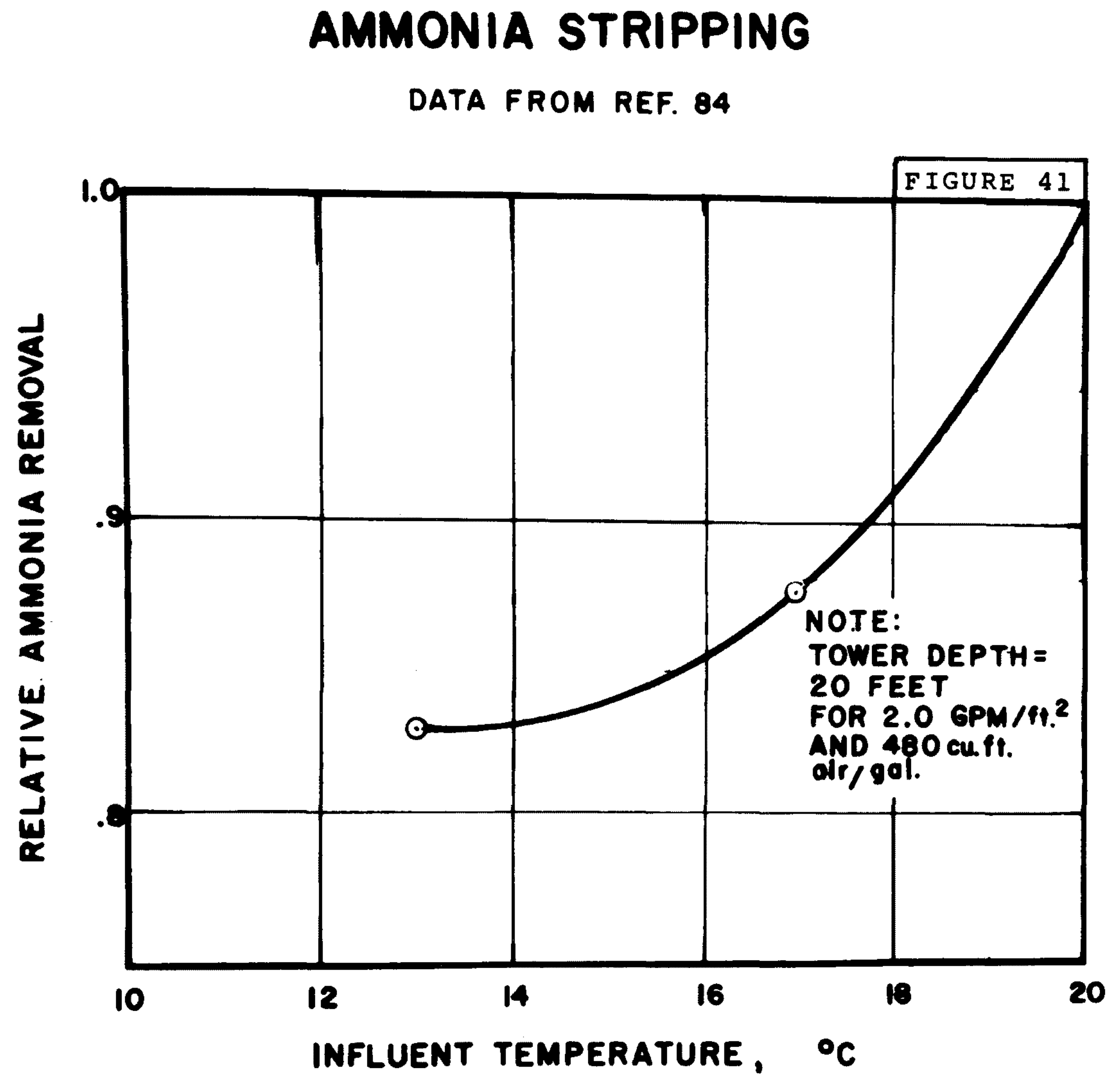


FIGURE 41 EFFECT OF TEMPERATURE ON AMMONIA STRIPPING EFFICIENCY FOR CONSTANT SIZE UNIT.

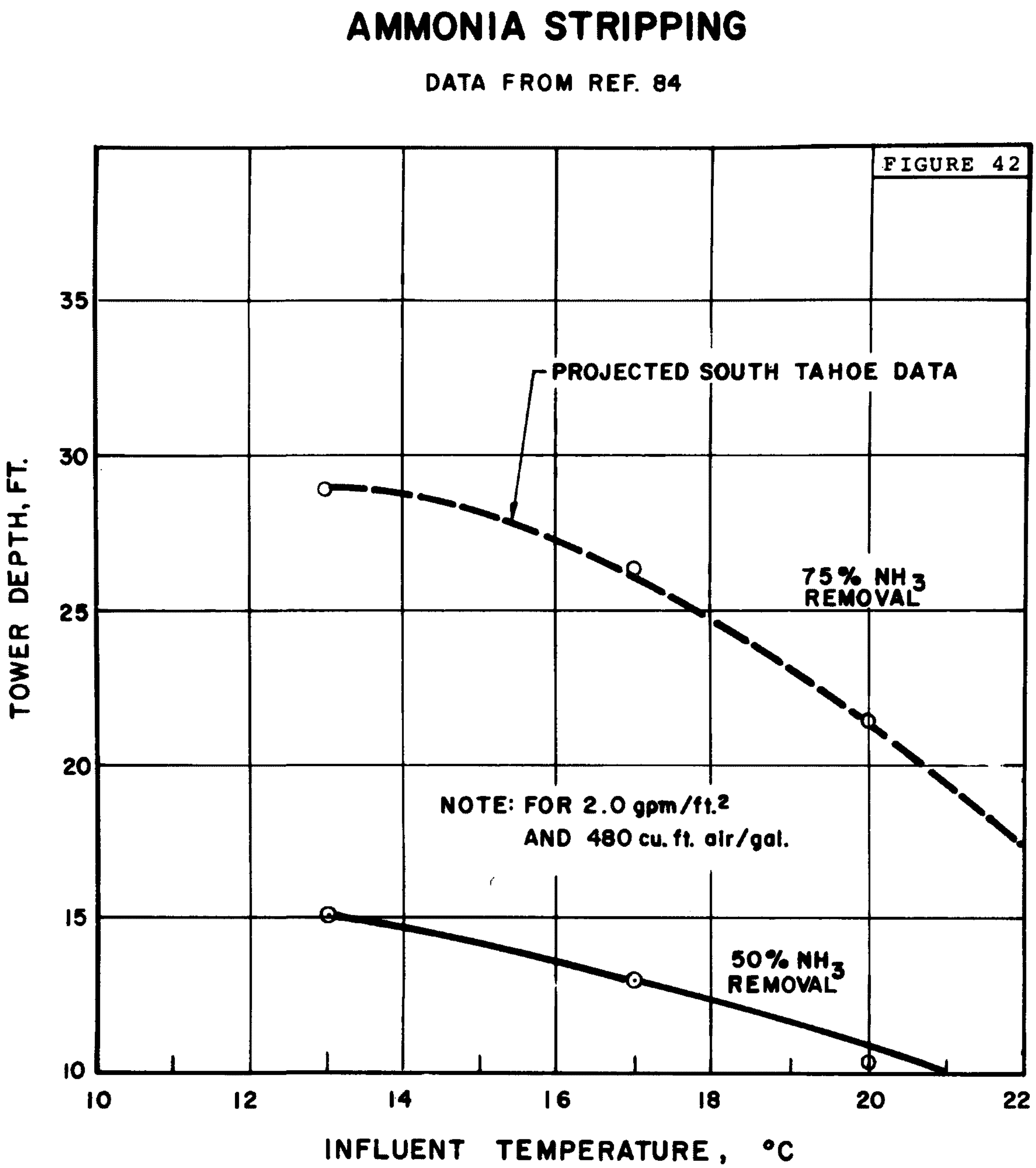


FIGURE 42 EFFECT OF TEMPERATURE ON AMMONIA STRIPPING UNIT SIZE FOR CONSTANT EFFICIENCIES.

Phosphorus Removal

Phosphorus removal may be achieved by "luxury" uptake in biological systems or by chemical precipitation. Other less commonly employed processes removing phosphorus include: electrodialysis, reverse osmosis, ion exchange, and distillation.

"Luxury" phosphorus removal in the activated sludge is achieved by incorporation of phosphorus in the synthesized cell mass beyond that amount required by the cell for synthesis on new cell material. Phosphorus may also be removed from recycled sludge by subjecting the sludge to controlled anaerobic or low pH conditions [105].

"Luxury" uptake of phosphorus in biological systems has not been refined to effectively predict its occurrence and extent based on design criteria and operational parameters [106]. However, it is predicted that phosphorus removal would increase at higher temperatures due to increased biological and chemical activities.

Chemical precipitation employing lime, alum, and ferric salts, or a combination thereof have been successfully employed for high phosphorus removals [105]. A survey of the solubilities of conventionally used chemical additives and resulting precipitates formed shows that the chemical additives are highly soluble and increase their solubility only slightly with increasing temperatures up to 60°C and that the precipitates formed are insoluble over this temperature range. Chemical precipitation of phosphates at higher temperatures should be more efficient due to enhanced physical separation and increased chemical reaction rates [108].

SECTION A-VII

REFERENCES

1. Carter, R. & Karassik, I., "Basic Factors in Centrifugal Pump Application," Water & Sewage Works, 21RP477.

2. Perry, J.H. & Perry, R.H., Engineering Manual, McGraw-Hill, 1959.

3. Rudolfs, W. et al., "Property of Sludge Which Effects Its Discharge Through 24-inch Pipe," Sewage Works Journal, Vol. 12, 1, 61, January 1940.

4. Brooke, M., "Flow Formula for Slurries," Chemical Engineering, January 1958.

5. Eckenfelder, W.W., Jr. & O'Connor, D.J., Biological Waste Treatment, Permagon Press, New York, 1961.

6. Fair, G.M. et al., Water and Wastewater Engineering, John Wiley, Vol. 2, 1968.

7. Chasick, A.H. & Theodore, B.B., "Using Graded Sand to Test Grit Removal Apparatus," JWPCF, Vol. 36, 7, 884, July 1964.

8. Renn, C.E., "Warm Water Effect on Municipal Supplies," JAWWA, Vol. 49, 405, 1957.

9. EPA, Process Design Manual for Suspended Solids Removal, EPA Technology Transfer, 1971.

10. Velz, C.J., "Influence of Temperature on Coagulation," Civil Engineering, Vol. 4, 345, 1934.

11. Camp, D.A., Root & Bhoota, B.V., "Effect of Temperature on Rate of Floc Formation," JAWWA, Vol. 32, 1913, 1940.

12. Willcomb, G.E., "Floc Formation and Mixing Basin Practice," JAWWA, Vol. 24, 1416, September 1932.

13. Parsons, W.A., Chemical Treatment of Sewage & Industrial Wastes, National Lime Association, 1965.

14. Stukenberg, J.R., "Biological-Chemical Wastewater Treatment," JWPCF, Vol. 43, 9, 1791, 1971.

15. Ives, K.J. & Sholji, I., "Research on Variables Affecting Filtration," Proc. ASCE, J. San. Eng. Div., Vol. 91, SA4, 1, 1965.

16. Lawrence, W.C., "Varying Wash Water Rates with Changes of Water Temperature," JAWWA, Vol. 22, 208, February 1930.

17. Vrablik, E.R., "Fundamental Principles of Dissolved Air Flotation of Industrial Wastes," Purdue Conference, XIV, 1959.

18. Burd, R.S., "A Study of Sludge Handling and Disposal," FWPCA, Pub. WP-20-4, 1968.

19. ASCE, WPCF, Sewage Treatment Plant Design, 1959.

20. Malina, J.F., "Temperature Effects on the Dewatering of Activated Sludge," Technical Report to U.S. PHS, EHE-69-02, CRWR-35.

21. Laboon, J.F., "Experimental Studies on the Concentration of Raw Sludge," S&IW, Vol. 24, 4, 423, April 1952.

22. Gassett, R.B. et al., "Influent of Various Parameters on Foam Separation," JWPCF, Vol. 37, 4, 470, April 1965.

23. Eldib, I.A. & Skororoski, R.M., "Renovation of Wastewater by Foam & Emulsion Fractionation," ESSO Res. & Eng. Co., Report No. PERD4M-63, 1963.

24. Rubin, E. et al., "Contaminant Removal from Sewage Plant Effluents by Foaming," FWQA, AWTR-5, 1963.

25. Eckenfelder, W.W., Jr., Industrial Waste Pollution Control, McGraw-Hill, 1969.

26. Jenkins, D. & McGauhey, P.H., "Broad Characterization of the Improvement of Sewage Effluents by Foam Separation," San. Eng. Res. Lab., U. of Calif., No. 65-5, 1965.

27. Miller, J.K.P. & Legatski, L.K., "Investigation of High Pressure Foam Wastewater Treatment Processes," Water Poll. Cont. Res. Ser., 17030ES 04/70, 1970.

28. Kennedy, R.R., "Thickening by Elutriation and Chemical Coagulation," in "Sludge Concentration, Filtration and Incineration," Continued Education Series No. 113, U. of Michigan, Ann Arbor, 1964, as quoted in Ref. 18.

29. Lamb, C., "Heat Treatment as an Aid to Sludge Dewatering - Ten Years' Full-Scale Operation," Journal Inst. of Sew. Purif., Part I, pp. 5-15, 1951.

30. Jepson, C. and Klein, L., "Heat Treatment of Sewage Sludge," Journal Inst. of Sew. Purif., Part I, pp. 36-45, 1951.

31. Teletzke, G.H., "Wet Air Oxidation of Sewage Sludge," A Paper Presented at the 16th University of Kansas Sanitary Engr. Conf., 1966.

32. Dick, R., "Role of Activated Sludge Final Settling Tanks," ASCE, JSED, April 1970.

33. Vesiland, P.A., "Estimation of Sludge Centrifuge Performance," ASCE, JSED, Vol. 96, SA3, 1970.

34. Alberton, O.E. & Guidi, E.E., "Centrifugation of Waste Sludges," JWPCF, Vol. 41, 4, 607, April 1969.

35. Landis, D.M., "Process Control of Centrifuge Operations," Chem. Eng. Prog., Vol. 66, 1, 51, January 1970.

36. Lubow, L.A., "Drying of Sludge on Heated Sludge Beds," Sewage Works Journal, Vol. 13, 5, 1011, May 1941.

37. Swanick, J.D. & Baskerville, R.C., "Sludge Dewatering on Drying Beds," Paper Presented at the London International Engineering Exhibition, April 1965.

38. Wiley, J.S., "Utilization and Disposal of Poultry Manure," Proc. 18th Ind. Waste Con., Purdue Univ., p. 515, 1963.

39. Corey, R.C., Principles and Practices of Incineration, Wiley, 1969.

40. Herman, E.R. & Gloyna, E.F., "Waste Stabilization Ponds, Formulation of Design Equations," J. Sewage & Ind. Wastes, August 1958.

41. Gloyna, E.F., Waste Stabilization Ponds, Vanderbilt, November 1971.

42. Barnhart, E.L., Aerated Lagoon Design, Vanderbilt, November 1971.

43. Gehm, H.W., "Activated Sludge at High Temperatures and High pH," Biological Treatment of Sewage and Industrial Wastes, McCabe & Eckenfelder Eds., Reinhold, New York, 1956.

44. Kshirsagar, S.R., "Effect of Temperature on Sewage Treatment," Instn. of Engineers (India), Vol. 44, 10, 3, June 1964.

45. Hunter, W.P. et al., "Temperature & Retention Time Relationships in Activated Sludge," Proc. 21st Ind. Waste Conf., Purdue Univ., EES #121, 1966.

46. Gotaas, H.B., "Effect of Temperature on Biochemical Oxidation of Sewage," SWJ, Vol. 20, 3, 441, May 1948.

47. Carpenter, W.L., "Temperature Relationships in Aerobic Treatment and Disposal of Pulp and Paper Wastes," JWPCF, Vol. 40, 5, 737, May 1968.

48. Rudolf, W. et al, "White Water Treatment, Aeration with Non-Flocculent Growths," JWPCF, Vol. 25, 1, 70, January 1953.

49. National Council for Stream Improvement, "Temperature Relationships in Aerobic Treatment and Disposal of Pulp and Paper Waste," Tech. Bulletin #191.

50. Gellman, I., "Influence of High Temperature on Activated Sludge," Johns Hopkins Univ., Res. Project, unpublished paper.

51. Maly, F., "Investigation of Biological Purification of Sewage on Aerobic Thermophilic Basis," M.S. Thesis for Reinisch-Westfalisch Technische Hochschule, Aachen (W. Ger.), Br. W.P. Abs., Vol. 40, 1965.

52. Dougherty, M.H., "Elevated Temperature Effect on Citrus Waste Activated Sludge," JWPCF, Vol. 30, 10, 1263, October 1958.

53. Howland, W.E., "Effect of Temperature on Sewage Treatment Processes," Sewage and Ind. Wastes, Vol. 25, 2, 161, February 1953.

54. Quirk, Lawler & Matusky Engineers, Technical Files #105, 175.

55. Velz, C.J., "A Basic Law for the Performance of Biological Beds," SWJ, Vol. 20, 4, 607, July 1948.

56. Committee Report, "Sewage Treatment at Military Installations," SWJ, Vol. 18, 5, p. 791, September 1946.

57. Rankin, R.S., "Evaluation of the Performance of Biofiltration Plants," ASCE, JSED, Vol. 120, 823, 1955.

58. Howland, W.E., "Flow Over Porous Media as in a Trickling Filter," Proc. 12th Ind. Waste Conf., Purdue Univ., 435, 1957.

59. Galler, S.W. & Gotaas, H.B., "Analysis of Biological Filter Variables," ASCE, JSED, Vol. 90, SA6, 59, 1964.

60. Weng, C. N., "Biological Fixed-Film Rotating Disks for Wastewater Treatment," Ph.D. Dissertation, New York University, New York, 1972.

61. Dague, R.L., "Anaerobic Activated Sludge," JWPCF, Vol. 38, 2, 220, February 1966.

62. Golueke, C.G., "Temperature Effects on Anaerobic Digestion of Sludge," Sew. & Ind. Wastes, Vol. 30, 1225, 1958.

63. Lawrence, A.L., & McCarty, P.L., "Kinetics of Methane Fermentation," JWPCF, Vol. 41, 2, R1, February 1969.

64. Fair, G.M. & Moore, E.W., "Heat and Energy Relationships in Digestion of Sewage Solids," Sewage Works Journal, Vol. 4, 755, March 1932.

65. Fair, G.M. & Moore, E.W., "Observations on the Digestion of a Sewage Sludge Over a Wide Range of Temperature," Sewage Works Journal, Vol. 9, 3, 1937.

66. Henkelkian, H. et al., "Effect of Change of Temperature on Thermophilic Digestion," Sewage Works Journal, Vol. 20, 5, 806, May 1968.

67. Maly, J. & Fadrus, H., "Influence of Temperature on Anaerobic Digestion," JWPCF, Vol. 43, 4, 641, April 1971.

68. Fair, G.M. & Moore, E.W., "Time and Rate of Sludge Digestion and Their Variation with Temperature," Sewage Works Journal, Vol. 6, 3, 1934.

69. Coulter, R.G., "Environment for Anaerobic Destruction of Organic Material," Public Works, Vol. 84, 5, May 1953

70. Imhoff, K. & Fair, G.M., Sewage Treatment, Wiley & Sons, New York, 1956.

71. Fischer, A.J. et al., "Plant Scale Tests on Thermophilic Digestion," SWJ, V. 17, 4, 718, 1945.

72. Garber, W.F., "Plant Scale Studies in Thermophilic Digestion," Sewage & Ind. Wastes, Vol. 26, 1203, 1954.

73. Popova, N.W. & Bolotina, O.T., "The Present State of Town Sewage & The Trend in Research Work in the City of Moscow," Advances in Water Pollution Research, Vol. 2, McMillan Co., New York, 1964.

74. Rawn, A.W., "Sludge Digestion Temperature Control with Live Steam," Water Works and Sewage, Vol. 90, 6, 246, June 1943.

75. Drier, P.E. & Obma, C.A., "Aerobic Digestion of Solids," Walker Proc. Equip. Co. Bulletin No. 26-S-18194, January 1963.

76. Lawton, G.W. & Norman, J.D., "Aerobic Sludge Digestion Studies," JWPCF, Vol. 36, 4, 495, April 1964.

77. Drier, D.E., "Aerobic Digestion of Solids," Purdue Indus. Wastes Conf. 18, p. 123, 1963.

78. Loehr, R.C., "Aerobic Digestion - Factors Affecting Design," Paper Presented at the 9th Great Plains Sewer Works Design Conference, March 1965.

79. Woodley, R.A., "A Study of Aerobic Biochemical Oxidation of Primary Sewage Sludge at Mesophilic and Thermophilic Temperatures," Unpublished Master's Thesis, Purdue University, W. Lafayette, Ind., 1961.

80. Rhines, C.E., "Fundamental Principles of Sewage Chlorination," Purdue Industrial Waste Conference, 20, 673, 1965.

81. Tashnival, C.L. et al., "Effect of Temperature on B.O.D. of Chlorinated Sewage," Chem. Abs., 66, 7450, 1967.

82. AWWA, Water Quality & Treatment, McGraw-Hill, 1971.

83. Goldstein, L.J. et al, "Continuous Flow Water Pasteurizer for Small Supplies," JAWWA, Vol. 52, 247, 1960.

84. Culp, R.L. et al., Advanced Wastewater Treatment, Van Nostrand, Reinhold, 1971.

85. Thirumurth, J.D., "Ozone in Water Treatment & Wastewater Renovation," Water & Sew. Works, Vol. 115, R-106, 1968.

86. Ogden, M., "Ozonation Today," Ind. Water Eng., Vol. 7, 6, 36, June 1970.

87. Marsh, G.R. & Panula, G.F., "Ozonation in the BOD Reduction of Raw Domestic Sewage," W&SW, Vol. 112, 10, October 1965.

88. Stumm, W., "Ozone as a Disinfectant for Water Supply & Sewage," Presented at Meeting of San. Sect., Boston Soc. of Civil Engineers, 1957.

89. Evans, F.L. & Ryckman, D.W., "Ozonated Treatment of Wastes Containing ABS," Proc. of 18th Ind. Waste Conf., Purdue Univ., 1963

90. Anon., "O_2 & O_3-R_x for Pollution," Chem. Engr., Vol. 77, 4, 46, February 1970.

91. Mixon, F.O., "Filterability Index & Microscreener Design," JWPCF, Vol. 42, 11, 1944, November 1970.

92. McCarty, P.L., "Biological Denitrification of Wastewaters by Addition of Organic Materials," Proc. 24th Ind. Waste Conf., Purdue Univ., p. 1271, 1969.

93. Nebolsine, R., "High Rate Filtration of Secondary Treatment Plant Effluent," Paper Presented at the New York Water Pollution Control Federation Annual Meetings, January 1972.

94. Morris, J.C. & Weber, W.J., Jr., Adsorption of Biochemically Resistant Materials from Solution 1, U.S. Dept. of HEW, AWTR-9, 1964.

95. Eckenfelder, W.W., Jr., "A Design Procedure for Biological Nitrification and Denitrification," Chemical Engineering Progress Symposium, Vol. 63, 78, 230, 1967.

96. Downing, A.L. et al., "Population Dynamics," Third Intl. Conf. on Water Pollution Research, Munich, 1966.

97. Mulbarger, M.C., "Nitrification and Denitrification in Activated Sludge Systems," JWPCF, Vol. 43, 10, 2059, October 1971.

98. Metcalf & Eddy, Inc., "Nitrification & Denitrification Facilities," For EPA Design Seminar for Wastewater Treatment, February 1972.

99. Wild, H.E., et al., "Factors Affecting Nitrification Kinetics," JWPCF, Vol. 43, 9, 1845, September 1971.

100. Beer, C., "Discussion of Evaluation of Anaerobic Denitrification Processes," ASCE, JSED, Vol. 96, SA6, 1452, December 1970.

101. Jerris, J., Personal Communication, February 1972.

102. Gulf South Res. Inc., "Methanol Requirement and Temperature Effects in Wastewater Denitrification," Water Poll. Res. Ser., 17010DHT09/70, August 1970.

103. Smith, J.M. et al., "Nitrogen Removal from Municipal Wastewater by Columnar Denitrification," Paper at EPA Design Seminar for Wastewater Treatment Facilities, February 1972.

104. Roesler, J.F. et al., "Simulation of Ammonia Stripping from Wastewater," ASCE, JSED, Vol. 97, SA3, 8182, June 1971.

105. Nesbitt, J.B., "Phosphorus Removal - The State of the Art," JWPCF, Vol. 41, 5, 701, May 1969.

106. Bunch, R.L., "Factors Influencing Phosphorus Removal by Biological Treatment," Water-1970, AICE, Vol. 67, 90, 1971.

107. Mulbarger, M.C. & Shefflett, D.G., "Combined Biological and Chemical Treatment for Phosphorus Removal," Water-1970, AICE, 67, 107, 1971.

108. EPA, Phosphorus Removal Process Design Manual, Technology Transfer, 1971.

109. Helfgott, T. & Hunter, J.V., "The Removal of Phosphates from Wastewater by Electrodialysis," Water-1969, AICE, Vol. 65, 218, 1969.

110. Anon., "High Temperature Electrodialysis Without Heat Addition," Saline Water Conversion Rep. - 1965, U.S. Dept. Int., Office of Saline Water, 187, 1964.

111. Brunner, C.S., "Pilot Plant Experience in Demineralization of Secondary Effluent Using Electrodialysis," JWPCF, Vol. 39, 10, R1, October 1967.

112. Pruyn, K.T., "Mathematical Model of the Electrodialysis Process," Water Poll. Contr. Res. Ser., 17090FTA07/69, July 1969.

113. Stephan, D.G. & Schaffer, R.B., "Wastewater Treatment and Renovation Status of Process Development," JWPCF, Vol. 42, 3, 399, March 1970.

114. Sanders, B.H. & Paris, E.F., "Electrodialysis - Principles & Applications in the Chemical & Food Industry," Ionics, Inc., 41A, Watertown, Mass.

115. Anon., "Fouling - Resistant Membranes for Water Desalting by Electrodialysis," Chem. Eng., Vol. 79, 1, 38, January 1972.

116. Cohen, H., "The Use of Reverse Osmosis in the Treatment of Industrial Effluents," from Developments in Water Quality Research, Shuval, H.I. (Ed.), Ann Arbor Science Pub., 1971.

117. Shuval, H.I. (Ed.), Developments in Water Quality Research, Ann Arbor Science Publishers Inc., 1971.

118. Aerojet Gen. Corp., "Reverse Osmosis Renovation in Municipal Wastewater," FWQA, 14-12-184, December 1969.

119. Osmonics Inc., Reverse Osmosis Systems, No. 67107, UMC.

120. Schmitt, C.R., "Don't Overlook Ion Exchange," Water & Wastes Eng., December 1971.

121. Mueller, J., Personal Communication, Manhattan College, Bronx, New York.

122. Levendusky, J.A., "Progress Report on the Continuous Ion Exchange Process," Water-1969, AICE, Vol. 65, 1969.

123. Dow Chemical Co., "Nitrate Removal from Wastewaters by Ion Exchange," Water Poll. Contr. Res. Ser., 17010FSJ01/71.

124. Envirogenics Co., "Investigation of a New Phosphate Removal Process," Water Poll. Contr. Res. Ser., 17010DJA11/70, November 1970.

125. Stamberg, J.B., "EPA Research & Development Activities with Oxygen Aeration," EPA Design Seminar for Wastewater Treatment, February 1972.

SECTION A-VIII

GLOSSARY

TERMS

Coagulation	The process of agglomeration of small particles into larger particles through agitation with or without the aid of chemicals
BOD	Biochemical Oxygen Demand
COD	Chemical Oxygen Demand
Floc	A particle formed by smaller particles through coagulation, usually by chemical addition
Hindered settling	Settling of particles in a liquid medium that does not behave as a single particle because of the interaction of other near-field particles
Solids flow	Rate of downward passage of solids in a thickening unit process
Aerobic	In the presence of oxygen
Anaerobic	In the absence of oxygen
Mesophilic	Describes a group of microorganisms that thrive at a temperature range of about 30°C to 40°C
Thermophilic	Describes a group of microorganisms that thrive at a temperature range of about 40°C to 50°C
Digestion	Process where complex organic compounds are converted to methane and carbon dioxide gases by anaerobic decomposition by anaerobic and facultative anaerobic microorganisms. Can also be accomplished by aerobic bacteria, producing carbon dioxide and ammonia.

Pathogen	A microorganism that produces disease
Disinfection	The elimination of pathogenic and other microorganisms by chemical addition or other means
Tertiary treatment	Processes that are added to secondary waste treatment facilities to improve the quality of the effluent

SYMBOLS

°C	degrees centigrade
°F	degrees Fahrenheit
gm	grams
mg	milligrams
cm	centimeters
l	liters
ml	milliliters
ft	feet
SF	square feet
in.	inches
hr	hour
min	minutes
sec	seconds
lbs	pounds
ac-ft	acre-feet
mph	miles per hour
gpd	gallons per day
ppm	parts per million

MLSS	mixed liquor suspended solids
Btu	British thermal units

APPENDIX B

ENGINEERING AND ECONOMIC FACTORS AFFECTING THE THERMAL HEATING OF WASTEWATER

The concept of operating a wastewater treatment plant at an elevated temperature as part of an integrated utility complex provides two opportunities for economic benefit. First, cost savings are possible through a reduction in size of the more efficient heated plant, and second, heat from the power or distillation plants which otherwise would be wasted can be recovered with a potential saving through the elimination of rejection equipment. Against these savings must be considered the technical and economic feasibilities associated with the introduction of waste heat to the wastewater process - an evaluation which is dependent upon:

1. The energy requirements of the treatment process.

2. The available energy sources.

3. Mechanisms for transferring energy from an available source to the wastewater.

Specific considerations were given to the effect of thermal energy addition on the processes employed in wastewater treatment; the materials of construction required to obtain consistently reliable performance without excessive maintenance and use of redundant equipment; the heat transfer coefficients that can be expected for each potential approach; and the effect of surface fouling on these coefficients. All calculations are based on an integrated wastewater treatment plant utilizing conventional primary and secondary techniques to treat 50 MGD of raw wastewater with an annual average (design) inlet temperature of 65^{o}F.

ENERGY REQUIREMENTS

In order to determine the relationship between the cost for a plant to process 50 MGD of wastewater and the temperature at which the plant is operated, an overall heat balance for the plant was established, incorporating information developed with regard to the heat losses through each unit of the proposed plant for various seasons of the year. As was described in Section VIII, the aeration tank was determined to be the unit operation which most affected the overall plant cost. It was shown that for temperatures in the aeration tank of greater than 86°F, virtually no further savings in the unit operation cost were achievable. Heat losses in the system up to and including the aeration tank, were computed to be 7°F, requiring that the incoming wastewater be heated to 93°F in order to realize the 86°F temperature in the aeration tank.

The heat required to achieve a temperature of 86°F in the aeration tank is dependent upon the point of heat addition. If heat is added prior to the grit chamber and a seven degree temperature loss is assumed in and prior to the aeration tank, the heat required would be 4.85×10^8 Btu/hr. If the heat is added prior to the primary settler, a reduction of approximately 0.17×10^8 Btu/hr is possible in the heat required. After accounting for the heat addition attributed to the 0.27 MGD of filter supernate at 95°F, heat losses due to radiation, conduction and convection of the heated wastewater to the atmosphere and the surrounding area as well as the cooling effect of the addition of approximately 4 MGD of thickener supernate at 85°F, an average temperature loss across the primary settler of one degree is anticipated. Thus, an influent of 92°F and an effluent of 91°F is acceptable to maintain an 86°F average temperature in the aeration tank.

The third alternative point for heat addition is prior to the aeration tank, in which case 4.50×10^8 Btu/hr would be required.

The energy requirements for winter operation will increase by 8.7×10^7 Btu/hr since the wastewater temperature will be approximately 5°F lower than for the design case; however, the energy requirements for the summer operation will decrease by the same amount since the summer wastewater temperature is assumed to be 5°F above the design base.

Four considerations dictate the choice of the point for heat addition: (1) the cost associated with the quantity of heat required at each point; (2) the type of equipment used to transfer the heat; (3) the conditions which could cause equipment fouling and thus reduce the effective heat transfer coefficient; and (4) corrosion and/or erosion of the heat transfer equipment.

AVAILABLE HEAT SOURCES

Having determined the quantity of energy necessary to raise the temperature of the wastewater to the design temperature, it is next necessary not only to identify the sources of available heat in the integrated utility complex, but also to determine if the energy is not so diffuse as to prohibit attainment of the necessary wastewater temperature. A number of sources of energy are available for consideration; namely,

Prime Steam

The first available heat source which may be considered is prime steam, defined as steam produced specifically for heating purposes by fuel combustion or produced in a steam generator or reboiler associated with a primary or secondary steam system of a power generating station.

Current fuel costs applicable to intermediate size industrial boilers are in the range of 60-90¢/10^6 Btu and increasing annually. This cost alone precludes further consideration of a single purpose energy source for heating of the wastewater. To this cost would have to be added, of course, the cost of heat transfer equipment, environmental controls, and in the case of a nuclear plant, a reboiler to eliminate the possibility of carryover of radioactivity.

Process Steam

The second source of energy is low pressure or process steam. For this analysis, process steam is assumed to be available saturated at a temperature of 285°F from a reboiler driven by a turbogenerator steam supply. Since the saturation pressure of 285°F steam is approximately 53 psi (abs), either surface heat transfer or direct transfer through injection of live steam can be utilized. Moreover, a reboiler is provided in the integrated plant to supply heat for the distillation process; hence, only the incremental cost increase for the distillation plant reboiler would have to be considered. The cost allocation approach which results in the lowest charges for process steam is achieved by determining the cost of prime steam produced for turbogenerator use, then prorating the cost of extracted steam on the basis of available energy. These cost calculations are described in detail in Section VI, Nuclear Steam Supply, and result in a cost of 29¢/10^6 Btu for process steam at 285°F.

Product Water

A potentially attractive approach consists of utilizing wastewater in lieu of salt water as the cooling media for the product water. Product water from the base case distillation plant leaves at an average temperature of 104°F and must be cooled before it can be pumped into a reservoir or directly into a pipeline. The maximum design temperature for this type of discharge is 85°F to prevent the introduction of undesirable concentrations of corrosion products in the water distribution system. If a design temperature of 74°F to 75°F for the product water were selected, there would be sufficient heat available to raise the wastewater stream to a temperature of approximately 93°F from the annual average inlet temperature of 65°F.

To effect the transfer of heat from the product water to the wastewater, liquid/liquid plate and shell and tube heat exchangers were considered. The capital cost of these heat exchangers is directly proportional to the surface area requirements and to the materials of construction.

Assuming that the product water is to be cooled from 104°F to 75°F and the wastewater is to be heated from 65°F to 92°F a log mean temperature difference (LMTD) of 11 is possible. Though this appears to be low, it is only slightly lower than the value calculated for the proposed integrated facility product water cooler using seawater as the coolant (LMTD of 12). The 92°F maximum wastewater temperature reflects the addition of heat after the grit chamber, to preclude possible excessive erosion of the heat exchanger by the raw wastewater.

The transfer of heat by liquid/liquid or steam/liquid surface heat exchange equipment is governed primarily by the temperatures at which the exchange takes place and by the resistances to heat transfer. In general, the heat transferred may be expressed at $Q = UA\overline{\Delta t}$ where Q is the heat exchanger duty in Btu/hr, U is the overall coefficient in Btu/hr/ft^2-$^\circ$F, A is the area of the heat exchange surface, and $\overline{\Delta t}$ is the log mean temperature difference. For liquid liquid heat exchangers such as a product water/screened and degritted wastewater interchanger, the overall resistance to heat transfer is caused by the following individual resistance:

(a) Cold fluid film resistance (R_{fc})

(b) Cold fluid fouling resistance (R_{sc})

(c) Metal tube wall resistance (R_w)

(d) Hot fluid fouling resistance (R_{sh})

(e) Hot fluid film resistance (R_{fh})

The total resistance, $R_t = R_{fc} + R_{sc} + R_w + R_{sh} + R_{fh}$.

U is related to R_t by the following:

$$U = \frac{1}{R_t} = (R_{fc} + R_{sc} + R_w + R_{sh} + R_{fh})^{-1}$$

Typical values of these resistances as derived for the product water/saltwater heat exchanger, assuming a velocity through the tubes of 3-4 feet per second, are as follows: (B-1, B-2)

R_{fc} = 0.0010

R_{sc} = 0.0005 (Seawater on inside of tubes)

R_w = 0.000068 (18 BWG 90/10 Cu-Ni 1" OD tube)

R_{sh} = 0.0005 (Product water on outside of tubes)

R_{fh} = 0.0010

R_t = 0.003068

The total of these resistances is 0.003068 and therefore,

$$U = \frac{1}{0.003068} = 326 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$$

A slightly more conservative value of 300 Btu/hr-ft^2-^{o}F was selected for the overall heat transfer coefficient for the product water/seawater heat exchanger used in the cost analysis.

Estimation of the fouling resistance, R_{sc}, the value for entering cold sewage stream has been attempted in various experimental tests; the Tubular Exchanger Manufacturer's Association (TEMA) has defined fouling coefficients (reciprocal of resistance) to be used for various types of coolants, temperatures, and velocities. For example, Kern (B-3) shows in his tabulation of these values, an average fouling resistance of 0.006 for Chicago Sanitary Canal coolant water treated for coolant use. Using this value in the previous example in place of the sum of R_{sc} + R_{sh} yields an overall resistance of 0.00717 or an overall heat transfer coefficient of approximately 140 Btu/hr-ft^2-^{o}F, thus the sensitivity of overall heat transfer coefficient to fouling resistance is readily apparent. On the other hand, based upon a computer synthesis of wastewater composition, a manufacturer estimated the fouling factor to be 0.0015, which would give rise to an overall heat transfer coefficient of 260 Btu/hr-ft^2-^{o}F (B-4).

Qualitative results from studies carried out using Contra Costa canal water (B-5) indicated that a fouling factor of approximately 0.0033 existed and that direct use of wastewater for cooling was attended by biogrowth and other fouling which had to be controlled by the addition of chlorine and other chemical additives.

An important approach to decreasing the resistance to heat transfer is to increase the velocity of the fluids. Three significant results of this strategy are:

1. The film resistance decreases (or, conversely, the film heat transfer coefficient increases).

2. Deposits, especially soft ones, are scrubbed off the surfaces more frequently, making the fouling resistance lower.

3. Pumping power increases due to increased frictional loss and the increased pass length required to maintain contact time.

The first of these results is apparent from Sieder and Tate's (B-6) equation for heat transfer

$$Nu = 0.023\ (Re)^{0.8}\ (Pr)^{0.33}\left(\frac{\mu_w}{\mu_b}\right)^{0.14}$$

where,

Nu = Nusselt number

Re = Reynolds number

Pr = Prandtl number

$\frac{\mu_w}{\mu_b}$ = Viscosity ratio of fluid at wall to that in bulk stream

The Nusselt number may be written as $\frac{hD}{k}$, the Prandtl number as $\frac{\mu C_p}{k}$, and the Reynolds number as $\frac{DV\rho}{\mu}$. Therefore Sieder and Tate's equation may be expressed as $h = ZV^{0.8}$ for given fluid conditions, where Z is a constant of proportionality. In the above formulae, the nomenclature used is:

h = film coefficient, Btu/hr-ft^2-^{o}F, or reciprocals of R_{fc} or R_{hf}.

D = tube diameter.

k = thermal conductivity of fluid.

μ = fluid viscosity.

ρ = fluid density.

C_p = specific heat.

V = fluid velocity.

Thus, it is possible to increase h by increasing velocity. Computation shows that doubling the velocity increases h by 1.73 times. Since $\frac{1}{h_{fc}} = R_{fc}$ and $\frac{1}{h_{fh}} = R_{fh}$ it can be shown that each film resistance could be reduced by 73 percent. Since the film resistance accounts only for a small fraction of the total resistance, the impact of such strategies appears to be low.

The rate at which deposits, whether biological or chemical, form and are removed from heat transfer surfaces, as a function of velocity, has not been clearly established.

In summary, it appears that liquid/liquid heat transfer rates will be governed primarily by the fouling resistances encountered in heating wastewater for enhancement of both primary and secondary treatment processes.

Based on the Chicago and the Contra Costa data, it is apparent that fouling resistances on the order of 0.003 to .005 can be expected with wastewater heating, resulting in overall heat transfer coefficients on the order of 140 to 193 Btu/hr-ft^2-^{o}F. However, based on the professional judgments of various heat exchangers and condenser manufacturers, higher values could be anticipated. Therefore, a comprehensive testing program would be in order to determine the appropriate fouling factors, the best method of cleaning and the required frequency of cleaning. For design purposes in this study, a wastewater fouling resistance of 0.0037 has been assumed; and the resulting overall heat transfer coefficient of 156 Btu/hr-ft^2-^{o}F was utilized for heat exchanger design calculation. Although this value may be conservative, the limited data available does not justify the assumption of a significantly higher coefficient.

Resistance values for the product water/wastewater exchange (assuming a velocity through the tubes of 3-4 fps, same as product water to saltwater case) are as follows:

R_{fc} = .00094

R_{sc} = .0037

R_w = .00025 (22 Bwg 304 SS 1" OD Tube)

R_{sh} = .0005

R_{fh} = $\underline{.0010}$

R_t = .00639

$$U = \frac{1}{R_t} = \frac{1}{.00639} = 156 \text{ Btu/hr-ft}^2\text{-}^o\text{F}$$

Turbogenerator Condenser

The incoming wastewater could utilize approximately 7 percent, or 4.85×10^8 Btu/hr of the turbogenerator heat (6.7×10^9 Btu/hr) and thus provide economic benefit for the integrated facility.

To effect the indirect transfer of heat from the turbogenerator exhaust steam to the wastewater, three types of surface condenser were considered: (1) standard type shell and tube, single and multipass condensers with the wastewater on the tube side; (2) a conceptual design of a box cooler with manifolded, singlepass, vertical tube sections, with the steam on the tube side. Vertical tubes are often preferred when the condensate must be appreciably subcooled below its condensation temperature; (3) sectionalized condenser. For these condensers, the cost is directly proportional to the surface area needed for condensation.

From Fourier's Law, the heat exchange area requirement (for a constant heat load) is inversely proportional to both the log-mean temperature difference (LMTD) and the overall heat transfer coefficient (U). Thus, cost is proportional to $1/U\overline{\Delta t}$. The major concerns with surface condensers are the unfavorable effects of fouling and corrosion, due to the use of wastewater, on the overall heat transfer coefficient. A decreased overall heat transfer coefficient results in high surface area requirements and the corrosive nature of wastewater suggests the use of more expensive construction materials.

The use of corrosion resistant materials for all heat transfer equipment is required on the bases of operational reliability and assurance of an uncontaminated product. Published results wherein carbon steel was used as heat transfer tubing in heating sanitary canal water show corrosion rates ranging from 6 mils per year (for canal water with corrosion inhabitators added) to 50 mils per year for untreated canal water. These tests were performed in recirculating tubes (B-7). Alternative materials of construction for heat exchangers include 90/10 and 73/30 Cu-Ni, USS 100 (stainless 409), stainless 316, stainless 304, titanium, monel, and brass and bronze alloys.

The requirements for reliability discussed above clearly exclude the use of carbon steel for the heat transfer surface.

Other studies (B-8) on polluted feedwaters indicate that the 70/30 Cu-Ni alloy may be superior to the 90/10 material in the presence of ammonia or hydrogen sulfide. However, further corrosion testing in the specific environment of interest appears to be a definite requirement in order to identify and measure the long term pitting, uniform corrosion rates and resistance to stress corrosion cracking.

Recent evidence (B-9, B-10, B-11) suggests that USS 100, Type 409 culvert grade stainless steel could have the necessary qualities of corrosion resistance, formability, and low cost to be considered for incorporation in a product water/wastewater heat exchanges. Preliminary data show the corrosion in saltwater to be only 2.4 mils for 4 years continuous service and 0.2 mils penetration for 6 years service in the Delaware and Monogahela Rivers. However, some deeper isolated pitting and crevice corrosion was noted. The cost for this material is 32-33¢/lb, as compared to approximately 60-70¢/lb for 304 stainless steel and 9-11¢/lb for carbon steel. Based on the discussions with fabricators of heat exchangers, the use of titanium is not prohibitively expensive at this time; however, the use of titanium or other "exotic" materials does not at the present time appear to be necessary to heat wastewater.

The design, operation, and maintenance of the wastewater and distillation plant is directed toward achieving the minimum overall costs consistent with such constraints as public safety and health. High availability by using low reliability equipment having poor corrosion resistance, may be achieved only with a great deal of redundancy in design and maintenance, or alternatively by using highly reliable, corrosion resistant, equipment with little redundancy or maintenance. Experimental analysis is required to determine the actual fouling and corrosion rates associated with various heat transfer surfaces.

The discussion of the relationship between fouling factors and overall heat transfer coefficients presented earlier in conjunction with liquid/liquid exchangers generally applies to liquid/vapor exchangers. However, liquid/vapor heat exchangers such as condensers or shell and tube type heat exchangers operating with cold wastewater and exhaust steam are expected to have somewhat higher overall heat transfer coefficients than would be achieved in liquid/liquid exchangers. Primarily, this is due to the fact that steam condensing resistances are lower than liquid film resistances (B-12). This advantage is not great since the primary resistance to heat transfer remains on the liquid side. For example, the overall resistance may be determined from the following individual resistances:

R_{fc} = 0.00094 (cold wastewater film resistance)

R_{sc} = 0.0037 (cold wastewater fouling resistance)

R_w = 0.00025 (22 BWG 304 SS, 1" OD tube)

R_{fh} = 0.0003 (condensing film resistance)

R_{sh} = 0.0002 (steam fouling resistance)

R_t = 0.00539

$$U = \frac{1}{R_t} = \frac{1}{0.00539} = 185 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}$$

The difference between R_{fh} and R_{sh} is the liquid/liquid case described earlier and the above values for the steam/liquid case is only .0005, which is only $\frac{(.0005)}{.00539} \times 100 = 9.3$ percent of the total resistance; consequently, improvements in the steam side coefficients cannot contribute greatly to the overall performance.

Again, manufacturers of heat transfer equipment have indicated that the fouling resistance of 0.0037 (based on Contra Costa experience), may be conservative, and have suggested values as low as 0.0005 (B-13). Such a favorable value would give an overall resistance of .00219 or an overall heat transfer coefficient of 456 Btu/hr-ft^2-$^\circ$F. Other manufacturers (B-4) have suggested a slightly more conservative overall heat transfer coefficient of 320 Btu/hr-ft^2-$^\circ$F.

For this application of vapor/liquid exchangers, it is particularly important to ensure that any leakage is in the direction of the vapor phase, which could be expected to be contaminated, albeit slightly, with radioactivity.

A relatively high head loss through the sectionalized condenser is possible, which coupled with the vacuum pump requirements for the vertical tube unit, are important considerations in using this alternative. However, manufacturers (B-13) have indicated that a pressure drop through the wastewater section of a condenser would be approximately 15-20 feet of head.

In summary, heat transfer manufacturers feel the tube fouling caused by the use of wastewater as a coolant in lieu of seawater, assuming that the wastewater has undergone filtration for the removal of large suspended solids, will require only minor changes in the condenser surface area requirement and maintenance and operating procedures with the additional installed cost involving approximately $650,000-$700,000 (B-13).

However, in order to reflect the absence of operating experience, this alternative was deferred until such time as experimental data could be developed to verify these opinions.

Distillation Plant Condenser

If a distillation plant matching the size of a wastewater treatment plant is utilized in conjunction with the wastewater treatment and power generating facilities, heat is available from the distillation plant condenser as a potential source for wastewater heat. Approximately 5 percent of the product water produced by the distillation plant comes from the last effect in the form of low pressure steam that must be condensed, and has a heat value of 12.5×10^8 Btu/hr. In order to heat the wastewater to $93^{\circ}F$ the temperatures of the steam formed in the last effect (No. 19) was increased from $91^{\circ}F$ at 1.5" Hg to $100^{\circ}F$ at 2" Hg by decreasing the overall distillation plant Δt.

The transfer of heat could be accomplished either by a sectionalized surface condenser or by direct injection of steam into the wastewater. If a sectionalized surface condenser is used to transfer this heat to the wastewater, the technical considerations are similar to those discussed in conjunction with the transfer of heat from the turbogenerator condenser. Again, the most important consideration is that of the extremely stringent control which must be exercised in design and operation to prevent contamination of the condensate.

Direct transfer of heat from the steam to the wastewater may be affected by either steam injection or barometric condensers. Usually, steam of sufficiently high pressure (> 15 psia) is used for direct injection or sparging, requiring the use of process steam, which has been shown to involve relatively high costs.

Alternatively, low pressure steam, as is available from the distillation plant, can be used but power must be supplied to the system to provide the driving forces necessary for getting the steam into the water since the water pressure at the bottom of a 10-foot high tank would be 16 times greater than the steam pressure. This injection could be accomplished by an eductor or ejector; however, these devices have high head losses, due to the velocities needed to create a suction of less than 2.5" Hg, necessary for the steam to be transferred.

Probably no other process equipment gives as much performance in terms of heat transfer per unit of investment cost, as the barometric condenser, which
inexpensive, provides direct contact between condensing vapor and cooling water without resistance of an intervening wall.

Because there is direct contact between the vapor and the cooling water, barometric condensers are used only where the condensable materials are not to be recovered. However, this still leaves a wide field for application where process steam is to be condensed. For example, barometric condensers are universally used to condense the steam vaporized from vacuum pans and from the last effect of multiple-effect evaporators, They are also used between stages, after the last stage of multistage steam ejectors, and on vacuum distilling columns to condense the process steam.

The use of barometric condensers ordinarily is not accompanied by the deposition of scale or other materials which impede the transfer of heat. Primarily this is due to the method by which heat is transferred - warm steam condenses directly on particles or flowing sheets of water without any intervening metal surface. Little recorded experience on the the use of raw, screened and degritted, primary settled, or secondary settled sewage streams is available since such coolants ordinarily are not used in power or industrial operations. However, it is anticipated that such streams can be heated in this equipment without impairment of the heat transfer function providing certain minimum maintenance procedures are carried out to prevent accumulation of excessive deposits. The growth of anaerobic organisms in the barometric condensers is expected since adequate concentration of nutrients and favorable temperatures exist. These micro-organisms are expected to be anaerobic since the low absolute pressure attained in the barometric condenser will result in the removal of most of the dissolved oxygen present in the waste stream. The degree to which such growth takes place is not known. Scaling by the deposition of calcium carbonate is not expected on the basis of Langelier Index values ranging from -0.35 to 0.00 for a sewage plant effluent in Bay Park, Long Island.

With this alternative, the condensate is not recovered but is recycled with the wastewater. This type of system has a lower capital and operating cost than the surface condenser approach; however, with this system, the steam that is added to heat the wastewater requires increased capacity of the equipment downstream of the heat addition point. As a result, the cost of the wastewater treatment facility as well as the distillation plant; increases, both in capital and operating costs. Since the total heat requirement of 4.85×10^8 Btu/hr is only half the available latent heat (12.5×10^8 Btu/hr) from the distillation plant overheads, two condensers, one for heating the wastewater, by means of barometric leg condensers and another through which seawater coolant was passed, would be required. The economics of this approach are discussed in a subsequent portion of this Appendix.

OVERALL COST OF HEAT ADDITION

The marginal costs of adding heat to the wastewater have been calculated for the various alternative methods previously described. In order to compare these costs on a common basis, the costs for a nonthermally-enhanced wastewater treatment plant integrated with a nuclear power plant and wastewater distillation plant have been calculated. The difference between this nonthermally-enhanced cost and the thermally-enhanced cost is the cost of adding heat. This difference is expressed both as \$/yr and ¢/Kgal product water. It should be noted that the costs presented here do not take into account any of the costs associated with the wastewater treatment equipment itself, but only those costs for the distillation plant and heat transfer equipment required for thermal enhancement of the wastewater plant.

As stated above, the base case has been taken as a non-thermally-enhanced integrated plant. The costs that have been estimated include the capital plus operating charges for the distillation plant, steam reboiler, and product post treatment.

Included in the capital cost estimates are the costs of size increases in distillation plant, product water cooler and steam supply (reboiler), and credits through size reductions in the distillation plant product water condenser, the integrated plant intake and outfall structures, and the pump stations associated with the hypothesized heat exchanger and/or barometric condensers. Operating cost calculations include estimates of labor, electric power, steam, chemical additions, and other operational and maintenance charges which are either directly associated with the added heat transfer equipment or are conventionally treated as a function of plant design capacities.

The following equations were developed during the course of this study to determine the cost allocations for the various unit processes within the plant complex.

Distillation Plant

The distillation plant cost has two major components:

(a) Heat transfer surfaces

(b) Volumetric containment and handling

Based upon various economic and design studies for 50 MGD desalination plants that had been prepared for the OSW, approximately 70 percent of the total plant cost is allocated to heat transfer surface and the remaining 30 percent to fluid handling and containment structures.

According to Fourier's Law, it can be shown that the heat transfer area is inversely proportional to the log mean temperature difference. On the basis of these facts and assumptions, the following equation was derived for determining the cost of the distillation plant as a whole:

$$C_{new} = C_{old} \left[0.03 + 0.7 \left(\frac{\Delta t_{old}}{\Delta t_{new}} \right)^{0.85} \right] \left[\frac{MGD_{new}}{MGD_{old}} \right]^{0.85}$$

For plants with constant Δt (driving force), the term $(0.3 + 0.7\ (\Delta t_{old}/\Delta t_{new})^{0.85}) = 1$. This is true because for the same Δt, the area does not change. For plants with lower Δt's, the area must increase by the ratio of $\Delta t/_{old}/\ \Delta t_{new}$ because for a lower Δt and the same heat transfer, A must increase according to Fourier's Law. The costs of the components (surface and volume) of the distillation plant are estimated to scale to the 0.85 exponential power.

C_{old} = \$59.375 x 10^6, best current estimate for 47.5 MGD product plant (B-15)

Δt_{old} = steam to first effect - steam from last effect = 285 - 100 = 185°F

MGD_{old} = 47.5 total product water

For the cases using a barometric condenser to inject steam for wastewater heating:

$$MGD_{new} = MGD_{old} + W_{exhaust}$$

where:

$W_{exhaust}$ = exhaust steam flow to the wastewater plant, expressed in millions of gallons per day

Steam Reboiler

The reboiler cost is estimated at \$2.5 x 10^6 for the 65°F reference case. For all other cases, the cost of this piece of equipment is assumed to vary according to the following equation:

$$C_{new} = C_{old} \left(\frac{Q_{new}}{Q_{old}} \right)^{0.85}$$

where:

C_{old} = \$2.5 x 10^6

Q_{new} = the steam load to the distillation plant, Btu/hr

Q_{old} = the steam load to the distillation plant for the 65°F case = 1.177 x 10^9 Btu/hr (constant)

Distillation Plant Condenser Credits

The condenser credit is based on the change in area requirements for varying steam loads. For the 65°F reference case, $A_o \approx 200{,}000\ ft^2$. The cost/$ft^2$ of surface has been estimated at \$15, installed. The following equation was used to obtain the credit for reduced surface requirements.

$$\text{Condenser credit } \Delta c = (15)\left(\frac{\left(W_c - W_{exhaust}\right)\left(\lambda_c\right)}{\left(\overline{\Delta t}\right)\ \left(U_o\right)} - \left(2 \times 10^5\right)\right)$$

where:

Δc = change in capital cost, \$ (This term is negative)

W_c = total steam from the last effect

$W_{exhaust}$ = exhaust steam to the wastewater plant

λ_c = latent heat of steam from last effect

$\overline{\Delta t}$ = log mean temperature difference of the condenser

U_o = overall heat transfer coefficient for the condenser = 550 Btu/hr-ft^2-°F (constant)

Intake/Outfall Structure Credit

The following expression was used to obtain the credits for a smaller intake/outfall structure due to decreased cooling water requirements:

$$\Delta I/O \text{ cost} = C_{old}\left[\left(\frac{MGD_{new}}{MGD_{old}}\right)^{0.6} - 1\right]$$

where:

$\Delta I/O$ = change in cost for the Intake and Outfall structures (This term is negative)

C_{old} = $\$1.22 \times 10^7$

MGD_{old} = 1070 MGD (constant)

MGD_{new} = new cooling water requirements

Product Water Cooler Debit

For the cases where the last effect temperature was raised in order to provide higher temperature steam to the wastewater plant, the product water temperature was also raised in order to keep the distillation plant in heat balance. Thus, a larger product cooler is needed. For these cases only, the following expression was used to obtain the debit:

$$\Delta PC = \left[\frac{W_p \left(\Delta t_{new} - \Delta t_{old} \right) Cs}{(U_o) \; (\overline{\Delta t})} \right]^{0.7}$$

where:

ΔPC = change in cost of the cooler, \$ (This term is positive and is added to the plant capital cost.)

W_p = product water flow in lb/hr = 16.5×10^6 (constant)

Δt_{new} = product water Δt across the cooler - varies according to case

Δt_{old} = product water Δt across the cooler for the 65^oF case = 20^oF (constant)

U_o = overall heat transfer coefficient for the cooler = 300 $Btu/hr\text{-}ft^2\text{-}^oF$ (constant)

$\overline{\Delta t}$ = log mean temperature difference for the product cooler = 12.35^oF (constant)

C_s = installed cost of heat transfer surface, $\$15/ft^2$ (constant)

If a liquid/liquid heat exchanger is used to transfer heat from the product water to the wastewater, from the cost of such an exchanger may be subtracted the cost of the product water cooler in the design plant.

Condenser Seawater Coolant Pump Power Credit

For the cases requiring a smaller condenser, less seawater is required for cooling purposes and, consequently, less electrical power. The following expression is based upon a 25 psi pressure drop for the seawater, a 75 percent pump efficiency, power cost at 9.1 mills/Kw-hr, and a 365-day operation.

$$\Delta CSCPP = \frac{(MGD_{new} - MGD_{old})\ (10^6)(25)(2.31)(0.7457)(8760 \times 9.1 \times 10^{-3})}{(1440)\ (3960)\ (0.75)\ (10^6)}$$

or

$$\Delta CSCPP = 0.00080268\ (MGD_{new} - MGD_{old})$$

where:

$\Delta CSCPP$ = change in power cost, $\$10^6$ (This term is negative.)

MGD_{old} = cooling water requirements for the 65°F case = 221 MGD (constant)

MGD_{new} = cooling water requirements for the smaller condensers and varies according to case selected

Product Cooler Seawater Pump Power Debit

The equation for the condenser seawater coolant pump is also used for the product cooler seawater pump except that MGD_{old} = 38.2 MGD. The results of these computation are positive and added to the annual operating and maintenance charges.

Calculated Cost

The dominant component of the cost of direct heat addition with process steam produced either by a separate fuel source or in a reboiler driven by extracted steam from the hypothesized power generating facilities is the cost of fuel. Considering a 95°F - single addition point case, 5.2×10^8 Btu/hr are required for 50 MGD wastewater feed; i.e., the heat requirement is approximately 0.25 x 10^6 Btu/Kgal. A fossil fuel cost of 60-90¢/10^6 Btu leads to an enhancement cost of 15-23¢/Kgal for fuel alone; i.e., without including the cost of heat transfer equipment, pumping power, O&M, etc. A process steam cost of 29¢/10^6 Btu is equivalent to an energy cost of 7.25¢/Kgal, exclusive of those costs associated with delivery of the energy. These partial costs, when considered in the perspective of total conventional wastewater treatment cost of 15-20¢/Kgal, preclude further consideration of the high temperature steam approach to thermal enhancement of the wastewater processes.

Three alternative ways of adding heat to the wastewater plant using barometric leg condensers have been costed for operation for 86°F, 95°F and 104°F temperature levels. These alternatives include:

(a) distillation plant exhaust steam-decreasing distillation plant Δt.

(b) distillation plant exhaust steam-constant distillation plant Δt.

(c) distillation plant exhaust steam combined with inter-effect extracted steam.

For alternative (a), as the design temperature of the wastewater treatment plant is increased, the final exhaust temperature from the distillation plant also increases, thus decreasing the overall distillation plant Δt since the steam temperature to the first distillation effect is constant. Because of this decreasing Δt and because of the increased throughput (due to the use of the barometric leg) the distillation plant capital cost will increase. Because of the increased throughput, the steam to the first effect increases, thus increasing the reboiler size and cost. Since the barometric condenser uses exhaust steam from the last effect, the condenser size (and cost) decreases due to the decreased loading. As the last effect temperature is increased, the product water temperature increases, thus requiring a larger and more costly product cooler. Associated with the decreasing condenser and increasing cooler is a decreasing cooling water load and decreasing intake/outfall structure cost. The cost allocation methods have previously been described. The costs for this alternative are shown in Table B-I and line 1 of the composite graph (Figure B-I).

Alternative (b) is essentially the same as (a) with the exception that the distillation plant Δt is held constant by increasing the steam temperatures to the first effect. All other increases and decreases in equipment size and cost are included where applicable. The costs for this alternative are shown in Table B-II and line 2 of the composite graph.

Alternative (c) combines the use of exhaust steam with steam that is extracted between distillation plant effects. In doing so, the plant Δt and final exhaust temperature remain constant for all levels of heating. As a result, the product cooler remains constant for all temperature levels. All other credits and debits have been estimated as previously described. The costs associated with this alternative are shown in Table B-III and line 3 of the composite graph.

All three of the above methods were calculated for two point heating; that is, heat has been added at the influent to the primary settling tank and at the influent to the final aeration tank with the warm recycle streams. This concept was originally examined when heat was assumed to be available at no cost.

Alternative (a) was also calculated using single point heating at a location ahead of the grit chamber and the primary settling tank only, without the warm recycle steam. These costs are shown in Table B-IV and line 4 of the composite graph.

As an alternative to the use of a barometric leg condenser system, the costs associated with a shell and tube liquid-liquid heat exchanger, transferring product water heat to the wastewater were examined. In this application, the product water is cooled from 104°F to 75°F, and the wastewater is heated from 65°F to 92°F (the latter temperature reflects the addition of heat after the wastewater has passed through the grit chamber, to preclude possible excessive erosion of the heat exchanger by the raw wastewater). Under these conditions, the heat to be exchanged is 4.68×10^8 Btu/hr, at a LMTD of 11.

TABLE B-I

BAROMETRIC LEG TWO POINT THERMAL ENHANCEMENT ALTERNATIVE A

Process Parameters and Cost Centers	Case Number 1	2	3	4
Process Parameters				
Wastewater treatment plant reference, Temperature, °F	65	86	95	104
Distillation plant feed rate, MGD	50	51.16	51.70	52.56
Distillation plant product rate, MGD	47.5	47.5	47.5	47.5
Distillation plant feed temperature, °F	65	88	93	101
Exhaust steam to barometric condensers, Temperature, °F	--	91	100	109
Exhaust steam to barometric condensers, lb/hr	--	402,500	588,990	889,055
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Capital Costs ($$10^6$)				
1. Distillation plant	59.375	60.604	62.937	65.384
2. Reboiler	2.500	2.552	2.576	2.614
3. Product post-treatment	1.300	1.300	1.300	1.300
4. Barometric condensers	--	.126	.129	.189
5. Barometric pump stations	--	.206	.208	.209
Subtotal	63.175	64.788	67.150	69.696
6. Condenser credit (-)	--	-0.952	-1.724	-2.402
7. Intake/outfall credit (-)	--	-0.274	-0.556	-0.730
8. Product cooler debit (+)	--	0	+0.534	+1.135
Net Total Capital Costs ($$10^6$)	63.175	63.562	65.404	67.699
Annual Fixed Charges ($$10^6$)				
1. Distillation plant @ 7.823%	4.645	4.741	4.924	5.115
2. Reboiler @ 7.823%	0.196	0.200	0.201	0.204
3. Product post-treatment @ 7.823%	0.102	0.102	0.102	0.102
4. Barometric condensers @ 7.823%	--	0.010	0.010	0.015
5. Barometric pump stations @ 7.823%	--	0.016	0.016	0.016
Subtotal	4.943	5.069	5.253	5.452
6. Condenser credit @ 7.823% (-)	--	-0.074	-0.135	-0.188
7. Intake/outfall credit @ 7.823%(-)	--	-0.021	-0.043	-0.057
8. Product cooler debit @ 7.823% (+)	--	0	+0.042	+0.089
Net Total Annual Fixed Charges ($$10^6$)	4.943	4.974	5.117	5.296
Annual Operating and Maintenance Charges ($$10^6$)				
1. Distillation plant				
Labor	0.408	0.418	0.423	0.430
Electric power	0.789	0.808	0.816	0.831
Chemicals	0.694	0.711	0.719	0.731
Spare parts and maintenance materials	0.891	0.913	0.923	0.939
Steam	2.991	3.064	3.098	3.152
2. Reboiler				
Operation and maintenance	0.025	0.026	0.026	0.026
3. Product post-treatment				
Operation and maintenance	0.020	0.020	0.020	0.020
4. Barometric condensers and pump stations				
Operation and maintenance	--	0.003	0.003	0.004
Electric power	--	0.131	0.131	0.132
Subtotal	5.818	6.094	6.159	6.265
5. Condenser pump power credit (-)	--	-0.057	-0.116	-0.153
6. Product cooler pump power debit (+)	--	0	+0.005	+0.009
Net Total Annual Operating and Maintenance Charges ($$10^6$)	5.818	6.037	6.048	6.121
Grand Total Annual Charges ($$10^6$)	10.761	11.011	11.165	11.417
Cost of Heat Addition ($$10^6$/yr)	--	0.250	0.404	0.656
¢/Kgal Product	--	1.44	2.33	3.78

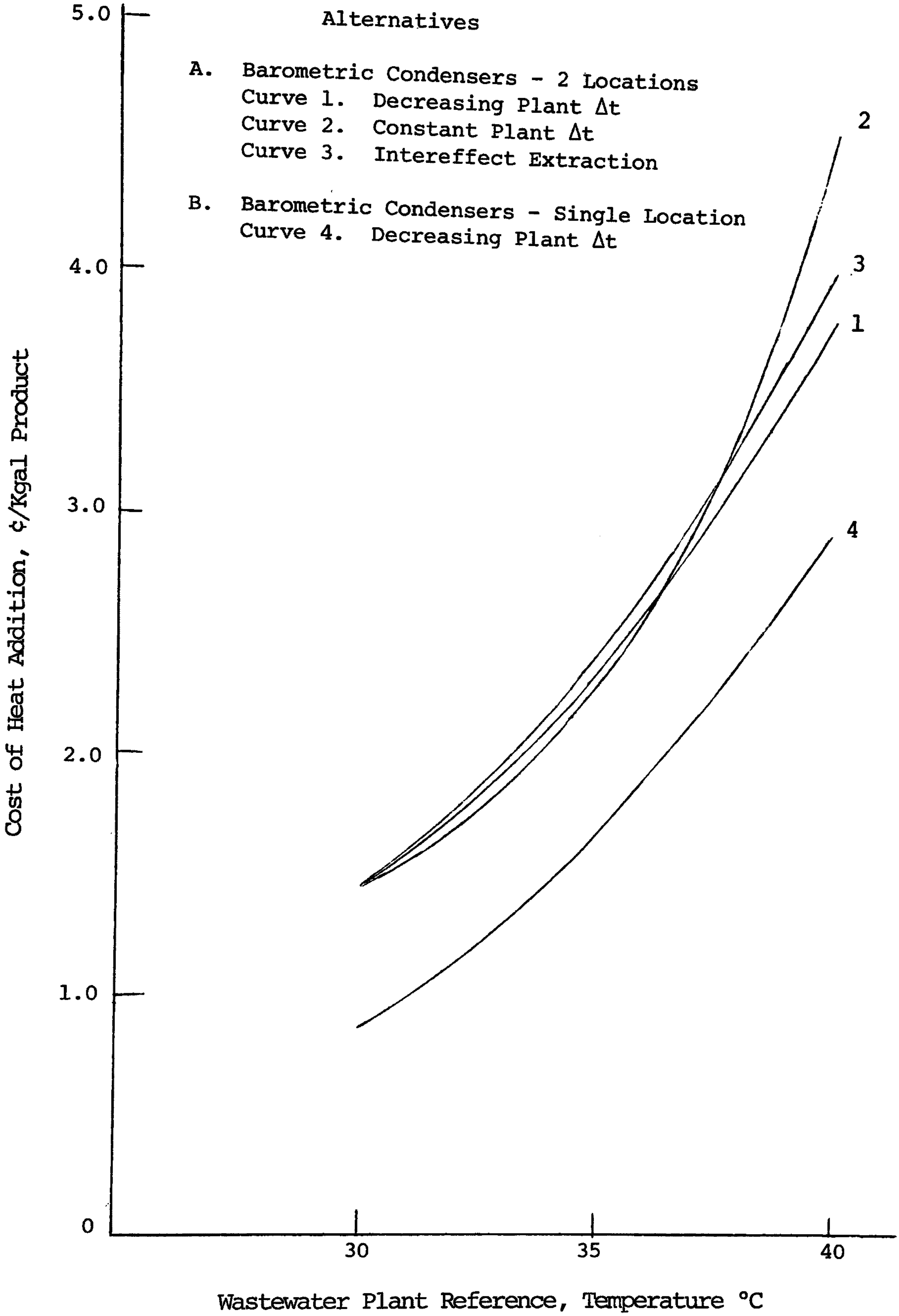


Figure B-I Cost Comparison of Thermal Enhancement Alternative Methods

TABLE B-II
BAROMETRIC LEG TWO POINT THERMAL ENHANCEMENT ALTERNATIVE B

Process Parameters and Cost Centers	Case Number 1	2	3	4
Process Parameters				
Wastewater treatment plant reference, Temperature, °F	65	86	95	104
Distillation plant feed rate, MGD	50	51.16	51.70	52.56
Distillation plant product rate, MGD	47.5	47.5	47.5	47.5
Distillation plant feed temperature, °F	65	85	93	101
Exhaust steam to barometric condensers, Temperature, °F	--	91	100	109
Exhaust steam to barometric condensers, lb/hr	--	402,500	588,990	889,055
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Capital Costs ($10⁶)	Same as Table I	Same as Table I		
1. Distillation plant			61.174	62.086
2. Reboiler			2.576	2.614
3. Product post-treatment			1.300	1.300
4. Barometric condensers			0.129	0.189
5. Barometric pump stations			0.208	0.209
Subtotal			65.387	66.398
6. Condenser credit (-)			-1.724	-2.402
7. Intake/outfall credit (-)			-0.556	-0.730
8. Product cooler debit (+)			+0.534	+1.135
Net Total Capital Costs ($10⁶)			63.641	64.401
Annual Fixed Charges ($10⁶)				
1. Distillation plant @ 7.823%			4.786	4.857
2. Reboiler @ 7.823%			0.201	0.204
3. Product post-treatment @ 7.823%			0.102	0.102
4. Barometric condensers @ 7.823%			0.010	0.015
5. Barometric pump stations @ 7.823%			0.016	0.016
Subtotal			5.091	5.170
6. Condenser credit @ 7.823% (-)			-0.135	-0.188
7. Intake/outfall credit @ 7.823%(-)			-0.043	-0.057
8. Product cooler debit @ 7.823% (+)			+0.042	+0.089
Net Total Annual Fixed Charges ($10⁶)			4.979	5.038
Annual Operating and Maintenance Charges ($10⁶)				
1. Distillation plant				
Labor			0.423	0.430
Electric power			0.816	0.831
Chemicals			0.719	0.731
Spare parts and maintenance materials			0.923	0.939
Steam			3.226	3.543
2. Reboiler				
Operation and maintenance			0.026	0.026
3. Product post-treatment				
Operation and maintenance			0.020	0.020
4. Barometric condensers and pump stations				
Operation and maintenance			0.003	0.004
Electric power			0.131	0.132
Subtotal			6.287	6.656
5. Condenser pump power credit (-)			-0.116	-0.153
6. Product cooler pump power debit (+)			+0.005	+0.009
Net Total Annual Operating and Maintenance Charges ($10⁶)			6.176	6.512
Grand Total Annual Charges ($10⁶)			11.155	11.550
Cost of Heat Addition ($10⁶/yr)			0.394	0.789
¢/Kgal Product			2.27	4.55

TABLE B-III

BAROMETRIC LEG TWO POINT THERMAL ENHANCEMENT ALTERNATIVE C

Process Parameters and Cost Centers	Case Number 1	2	3	4
Process Parameters				
Wastewater treatment plant reference Temperature, °F	65	86	95	104
Distillation plant feed rate, MGD	50	51.16	51.82	52.48
Distillation plant product rate, MGD	47.5	47.5	47.5	47.5
Distillation plant feed temperature, °F	65	88	93	101
Exhaust steam to barometric condensers, Temperature, °F	--	91	91	91
Exhaust steam to barometric condensers, lb/hr	--	402,500	347,860	347,860
Extracted steam to barometric condensers, Temperature, °F	--	--	105	105
Extracted steam to barometric condensers, lb/hr	--	--	280,345	387,285
Extracted steam to barometric condensers, Temperature, °F	--	--	--	119
Extracted steam to barometric condensers, lb/hr	--	--	--	125,010
Capital Costs ($10⁶)				
1. Distillation plant			62.350	65.277
2. Reboiler			2.581	2.610
3. Product post-treatment			1.300	1.300
4. Barometric condensers			0.184	0.240
5. Barometric pump stations			0.266	0.267
Subtotal			66.681	69.694
6. Condenser credit (-)			-1.608	-2.152
7. Intake/outfall credit (-)			-0.469	-0.638
8. Product cooler debit (+)			0	0
Net Total Capital Costs ($$10^6$)			64.604	66.904
Annual Fixed Charges ($$10^6$)				
1. Distillation plant @ 7.823%			4.878	5.107
2. Reboiler @ 7.823%			0.202	0.204
3. Product post-treatment @ 7.823%			0.102	0.102
4. Barometric condensers @ 7.823%			0.014	0.019
5. Barometric pump stations @ 7.823%			0.021	0.021
Subtotal			5.193	5.429
6. Condenser credit @ 7.823% (-)			-0.126	-0.168
7. Intake/outfall credit @ 7.823%(-)			-0.037	-0.050
8. Product cooler debit @ 7.823% (+)			0	0
Net Total Annual Fixed Charges ($$10^6$)			5.054	5.235
Annual Operating and Maintenance Charges ($$10^6$)				
1. Distillation plant	Same as Table I	Same as Table I		
Labor			0.424	0.429
Electric power			0.819	0.830
Chemicals			0.720	0.730
Spare parts and maintenance materials			0.925	0.937
Steam			3.104	3.147
2. Reboiler Operation and maintenance			0.026	0.026
3. Product post-treatment Operation and maintenance			0.020	0.020
4. Barometric condensers and pump stations Operation and maintenance			0.004	0.005
Electric power			0.174	0.218
Subtotal			6.216	6.342
5. Condenser pump power credit (-)			-0.095	-0.127
6. Product cooler pump power debit (+)			0	0
Net Total Annual Operating and Maintenance Charges ($$10^6$)			6.121	6.215
Grand Total Annual Charges ($$10^6$)			11.175	11.450
Cost of Heat Addition ($$10^6$/yr)			0.414	0.689
¢/Kgal Product			2.39	3.97

TABLE B-IV
BAROMETRIC LEG ONE POINT THERMAL ENHANCEMENT ALTERNATIVE A

Process Parameters and Cost Centers	Case Number 1	2	3	4
Process Parameters				
Wastewater treatment plant reference, Temperature, °F	65	86	95	104
Distillation plant feed rate, MGD	50	51.00	51.44	51.88
Distillation plant product rate, MGD	47.5	47.5	47.5	47.5
Distillation plant feed temperature, °F	65	80	86	93
Exhaust steam to barometric condensers, Temperature, °F	--	91	100	109
Exhaust steam to barometric condensers, lb/hr	--	347,860	499,425	652,505
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Extracted steam to barometric condensers, Temperature, °F	--	--	--	--
Extracted steam to barometric condensers, lb/hr	--	--	--	--
Capital Costs ($10^6)				
1. Distillation plant		60.436	62.658	65.073
2. Reboiler		2.544	2.564	2.584
3. Product post-treatment		1.300	1.300	1.300
4. Barometric condensers		0.116	0.123	0.141
5. Barometric pump stations		0.056	0.056	0.056
Subtotal		64.452	66.701	69.154
6. Condenser credit (-)		-0.824	-1.553	-2.030
7. Intake/outfall credit (-)		-0.235	-0.515	-0.650
8. Product cooler debit (+)		0	+0.534	+1.135
Net Total Capital Costs ($10^6)		63.393	65.167	67.609
Annual Fixed Charges ($10^6)				
1. Distillation plant @ 7.823%		4.728	4.902	5.091
2. Reboiler @ 7.823%		0.199	0.201	0.202
3. Product post-treatment @ 7.823%		0.102	0.102	0.102
4. Barometric condensers @ 7.823%		0.009	0.010	0.011
5. Barometric pump stations @ 7.823%		0.004	0.004	0.004
Subtotal		5.042	5.219	5.410
6. Condenser credit @ 7.823% (-)		-0.064	-0.121	-0.159
7. Intake/outfall credit @ 7.823%(-)		-0.018	-0.040	-0.051
8. Product cooler debit @ 7.823% (+)		0	+0.042	+0.089
Net Total Annual Fixed Charges ($10^6)		4.960	5.100	5.289
Annual Operating and Maintenance Charges ($10^6)				
1. Distillation plant				
Labor		0.417	0.420	0.424
Electric power	Same as Table I	0.806	0.813	0.820
Chemicals		0.709	0.715	0.721
Spare parts and maintenance materials		0.910	0.918	0.926
Steam		3.054	3.082	3.109
2. Reboiler				
Operation and maintenance		0.025	0.026	0.026
3. Product post-treatment				
Operation and maintenance		0.020	0.020	0.020
4. Barometric condensers and pump stations				
Operation and maintenance		0.002	0.002	0.002
Electric power		0.056	0.056	0.056
Subtotal		5.999	6.052	6.104
5. Condenser pump power credit (-)		-0.049	-0.108	-0.138
6. Product cooler pump power debit (+)		0	+0.005	+0.009
Net Total Annual Operating and Maintenance Charges ($10^6)		5.950	5.949	5.975
Grand Total Annual Charges ($10^6)		10.910	11.049	11.264
Cost of Heat Addition ($10^6/yr)		0.149	0.288	0.503
¢/Kgal Product		0.86	1.66	2.90

Potential savings which result from this arrangement include approximately $500,000 from the reduction in size of the intake and outfall structures (Table 10, Section IX), $1,355,000 for the product water cooler in the base case (Case II) and $1,497,000 in wastewater treatment plant costs due to heat addition. If these savings of $3,332,000 were to be entirely offset by the cost of a liquid/liquid heat exchanger, the maximum permissible area costs for installed equipment as a function of attainable overall heat transfer coefficients are shown in Table B-V.

TABLE B-V

BREAKEVEN COSTS FOR A PRODUCT WATER/WASTEWATER HEAT EXCHANGER AS A FUNCTION OF THE HEAT TRANSFER COEFFICIENT ATTAINABLE

U Btu/hr-Ft^2-^{o}F	A Sq Ft	\$/Ft^2 (Installed)
150	2.84×10^5	\$11.70
200	2.13×10^5	\$15.60
250	1.70×10^5	\$19.60
300	1.42×10^5	\$23.45
350	1.22×10^5	\$27.30
400	1.06×10^5	\$31.40

REFERENCES

B-1 Personal correspondence with Anthony J. Giuriceo, Sales Engineer, Mesco Tectonics, Inc.

B-2 "Standards of the TEMA", Tubular Exchanger Manufacturer's Association, 3rd Edition, New York, New York, 1952.

B-3 Kern, D.Q., Process Heat Transfer, McGraw-Hill Book Co., Inc., New York, 1950.

B-4 Personal correspondence with Roma Bhattacharya, Product Specialist Heat Transfer Dept., Bell & Gossett and with Walter C. Bosch, Wallace Eannace Associates, Inc., Bell & Gossett's, New York Representatives.

B-5 "Central Costa County Sanitary District and Contra Costa County Water District", Bechtel Corporation, Contra Costa, California. EPA Grant No. 17080 FSF.

B-6 Sieder, E.N., and G.E. Tate, Industrial Engineering Chemistry, No. 28, McGraw-Hill Book Co., New York, 1936.

B-7 Weddle, C.L., and H.N. Masri, "Industrial Use of Renovated Municipal Wastewater", Bechtel Corporation, San Francisco, California, March 1972.

B-8 Todd, Bailie, and Tuthill, "Desalination-Lower Cost Water by Proper Materials Selection", Desalination Conference, Yugoslavia, September 1970.

B-9 Preliminary corrosion performance data on stainless steels - U.S. Steel Research Laboratory.

B-10 Evaluating Material Performance in 3,000 GPD stainless steel desalination test plants - 18 months operation by H.H. Lawson & T.R. Harkin, Paper No. 25, 1972, Conference of the National Association of Corrosion Engineers.

B-11 R. Smith, Corrosion Resistence of Various Steels in Liquid Manure Proj. No. 57.001-003 (13) April 1969.

B-12 Rohsenow, Warren M., and Harry Choi, Heat, Mass and Momentum Transfer, Prentice-Hall, Inc., Englewood Cliffs, New Jersey, 1961.

B-13 Personnal correspondence with William R. Ryan, Sales Manager Water Products & Environmental Systems, Westinghouse Electric Corp. Power System, Heat Transfer Division.

APPENDIX B

ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS	MEANING
A	Area
Btu	British thermal unit
C_{new}	New cost
C_{old}	Old cost
Cp	Specific Heat
C_s	Cost of heat transfer surface
Cu-Ni	Copper Nickel
D	Tube diameter
^oF	Fahrenheit
h	Film coefficient
Hg	Mercury
hr	Hour
k	Thermal conductivity of fluid
Kgal	A thousand gallons
LMTD	Log mean temperature difference
MGD	Million gallons per day
Nu	Nusselt number
Pr	Prandtl number
Q	Heat exchange duty in Btu/hr
Q_{new}	Steam load to the distillation plant
Re	Reynolds number
R_{fc}	Cold fluid film resistance
R_{fh}	Hot fluid film resistance
R_{sc}	Cold fluid fouling resistance

ABBREVIATIONS (Cont'd)	MEANING
R_{sh}	Hot fluid fouling resistance
R_t	Total Resistance
R_w	Metal tube wall resistance
TEMA	Tubular Exchanger Manufacturer's Assoc.
U	Overall heat transfer coefficient Btu/hr-ft^2-^{o}F
USS	United States Steel
V	Fluid velocity
W_c	Total steam from last evaporator effect
$W_{exhaust}$	Exhaust steam flow to wastewater
W_p	Product water flow in lb/hr

GREEK SYMBOLS

Δ C	Change in capital cost
Δ CSCPP	Change in power cost
Δ I/O	Change in cost of intake and outfall structure
Δ PC	Change in cost of product cooler
$\overline{\Delta t}$	Log mean temperature differential
λ_c	Latest heat
ρ	Fluid density
μ	Fluid viscosity

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
EPA-670/2-74-080		

4. TITLE AND SUBTITLE	5. REPORT DATE
STUDY OF AN INTEGRATED POWER, WATER AND WASTEWATER UTILITY COMPLEX	December 1974; Issuing Date
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
New York State Atomic and Space Development Authority	
9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO.
New York State Atomic and Space Development Authority 230 Park Avenue New York, New York 10017	1BB043
	11. CONTRACT/~~GRANT~~ NO.
	17080 HHV
12. SPONSORING AGENCY NAME AND ADDRESS	13. TYPE OF REPORT AND PERIOD COVERED
National Environmental Research Center Office of Research & Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268	Final
	14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES

16. ABSTRACT

This study evaluates, technically and economically, a new approach to siting power generation, wastewater treatment and water supply facilities. It is included that the integrated facility results in more efficient utilization of land and water resources, produces a net reduction in undesirable process effluents, and achieves at a reduced cost many of the environmental quality goals sought today. In particular, the use of waste heat for the beneficiation of wastewater treatment was determined to be sufficiently promising to merit further investigatory research. The integrated facility studied will supply 1000 Mw of electric power at 9.1 mills/Kw-hr, will provide secondary treatment for 50 MGD of wastewater for 15¢/1000 gal., and will produce 47.5 MGD of high quality potable water for approximately 62¢/1000 gal. utilizing low quality steam and waste heat. A three-phase follow-on research and demonstration program is defined and is directed toward the development of the further design and performance information necessary to permit the undertaking of full scale integrated facilities. This report was submitted in fulfillment of Project Number 17080 HHV under the sponsorship of the Environmental Protection Agency.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Waste treatment, Water treatment, Waste water, Heat recovery, Cost effectiveness	Waste heat utilization, Wastewater treatment, *Wastewater distillation *Facility integration	13B

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(This Report)*	21. NO. OF PAGES
Release to Public	Unclassified	238
	20. SECURITY CLASS *(This page)*	22. PRICE
	Unclassified	

☆ U. S. GOVERNMENT PRINTING OFFICE: 1975-657-588/5326 Region No. 5-11