Evaluation of the Full-Scale Application of Anaerobic Sludge Digestion at the Blue Plains Wastewater Treatment Facility Washington, D.C.

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FINAL REPORT

Evaluation of the Full-Scale Application
of Anaerobic Sludge Digestion at the
Blue Plains Wastewater Treatment Facility
Washington, D.C.

Submitted to:

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SUMMARY

This study investigated the application of a new mesophilic-thermophilic anaerobic sludge digestion process to the existing mesophilic sludge digestion system at the District of Columbia Wastewater Treatment Facility (WWTP). As part of this study, improvements in the existing mesophilic digestion operation and possible application of thermophilic digestion technology were also evaluated. The study was prepared under Contract No. 68-01-5913 for the US Environmental Protection Agency, with the cooperation of the city of Washington DC and the Wastewater Treatment Facility's technical personnel. The analyses are presented in detail to facilitate the use of the approach as a case study model for other wastewater treatment facilities considering the application of the process.

The mesophilic-thermophilic digestion process is a new concept in the treatment of municipal wastewater treatment sludges. It involves a two-step digestion process, where the first step operates under mesophilic process conditions, (that is, digestion with anaerobic microorganisms that thrive at 90 to 100°F) and the second step operates under thermophilic process conditions (that is, digestion with anaerobic microorganisms that thrive at 120 to 130°F). The mesophilic process is the most commonly employed digestion process; the thermophilic process has had limited application in this country but is used regularly in the U.S.S.R. Use of the mesophilic-thermophilic process has been documented only at the Rockaway facility in New York City.

The development of standard process conditions for anaerobic sludge digestion is described in this report, including operating conditions and process effectiveness for the Rockaway treatment facility.

The development and application of the mesophilic-thermophilic process has been pioneered by Mr. Wilbur Torpey. A full-scale application and evaluation of the effectiveness of this dual process approach has been been undertaken by the Rockaway Pollution Control Plant in New York City. The results of the Rockaway trials indicate that the physical characteristics of mesophilic-thermophilic digested sludge are improved and that these can result in (1) a significant reduction in the dewatering requirements of the stabilized material, and (2) a more stabilized residual that is nearly pasteurized and inert. The savings from dewatering improvements appear to be significant and outweigh any increased costs associated with operating the two processes rather than either of the conventional single processes. Moreover, there is the possibility that use of the mesophilic-thermophilic process may result in an overall reduction in digestion time. This is because the mesophilic system tends to dampen any fluctuations in waste sludge characteristics, thus providing a consistent feed to the thermophilic system. As a result, the thermophilic system can be designed to operate with a reduced safety factor for feed sludge variation. Since the thermophilic system is a higher rate process than the mesophilic, a net reduction in the overall detention time may be achieved.

These considerations and the success of the Rockaway trial resulted in the desire to investigate the feasibility of applying the mesophilic-thermophilic process to a major wastewater treatment facility. The District of Columbia WWTF was selected because: 1) the influent wastewater was mainly domestic, as was the Rockaway influent; 2) the facility has anaerobic digesters in operation; 3) the facility will require modifications to upgrade the sludge handling facilities; and 4) the staff at the facility were technically capable and willing to cooperate with the investigation.

The investigations at the District of Columbia WWTF involved a detailed review of existing plant operations including the development of flow and mass balances using existing data developed for the sludge management area. It was necessary to develop a detailed piping schematic for the digester area, as this was not available. A tracer study was performed on the existing twelve anaerobic digesters to determine the useful volume. Interviews with plant staff were used to further establish the existing operating conditions. These sludge operating conditions were characterized according to the solids generated per million gallons of influent wastewater flow.

The analytic data were evaluated in conjunction with financial data to identify the cost of treatment options. Such an evaluation was performed to determine alternative costs for the following five operating conditions:

1) Current operating practices (January to June 1980), with anaerobic digestion treatment of a partial sludge stream.

2) Spring 1981, after completion of plant modifications in progress during the study, with anaerobic digestion treatment of a partial sludge stream.

3) Mesophilic digestion of all sludge.

4) Thermophilic digestion of all sludge.

5) Mesophilic-thermophilic digestion of all sludge.

The financial analyses were based on a mixed primary-secondary sludge, with sludge processed by gravity thickening, dissolved air flotation, anaerobic digestion, elutriation, vacuum filter dewatering, and final disposal through land application. Undigested sludge in excess of the existing anaerobic digestion system capacity is limed, dewatered and disposed of through composting or land trenching.

The engineering and financial evaluations of full-scale application of the mesophilic-thermophilic anaerobic digestion process concluded that: a limited expansion of digester capacity is required to handle the entire sludge stream; there would be digester gas available for sale to outside interests after internal heating requirements were satisfied; and the cost of sludge handling could be reduced by $24 to $31 per million gallons of influent flow. The analysis also indicated that the improved characteristics of the mesophilic-thermophilic digested sludge could reduce chemical conditioning requirements so cost would be almost $7 less (per million gallons of influent flow) than mesophilic digestion and almost $4 less than thermophilic digestion. Moreover, the Rockaway results indicate that additional savings may be incurred during disposal because the stabilized material should be appropriate for use as a soil conditioner.

RECOMMENDATION

Based on our review of the anaerobic sludge digestion options and how they could be adapted to the existing facilities, it is our strong recommendation that the thermophilic anaerobic digestion process is the best fit for the District of Columbia and should be implemented on a full scale basis. This recommendation is based on our thorough review of the present state of practice within the United States and other countries. We heavily weighed the 30 years of successful experience at the City of Los Angeles Hyperion Plant and their decision to convert the total digestion system to the thermophilic process. Another factor was the successful conversion, over 2 1/2 years ago, of the entire mesophilic digestion system to the thermophilic mode by the city of Denver. The City of Chicago has also shown that the capacity of a mesophilic digestion tank can be doubled by converting to thermophilic operation.

The thermophilic process is the option that could be initiated with a minimum of time and money. Other significant advantages of the process are (a) increased sludge processing capability, (b) improved sludge dewatering as to coagulant demand and yield, and (c) increased destruction of pathogens, all of which are pertinent to the needs of the District of Columbia Treatment Facility.

It is especially important that, prior to start up, an independent engineer check the structural competency of the existing digesters and piping at the thermophilic temperatures, as well as the temperature control system.

We strongly recommend and urge that the transition from mesophilic to thermophilic operation be implemented as rapidly as possible in order to alleviate the solids handling problems with the metropolitan area. A carefully formulated plan for the transition should be prepared so that the transition be carried out with a minimum of interference with plant operations.

CONCLUSIONS

1. It has been found possible to process the total waste sludge flow using the existing facilities for the thermophilic anaerobic digestion process. In that connection the accumulated grit in the digestor tanks will not have to be removed.

2. Adoption of the thermophilic digestion option requires the least capital expenditure, would be the most expedient solution to the sludge management problem and would yield substantial operational cost savings. It also offers the potential, because of the pathogen kill, of eliminating the need for composting.

3. The mesophilic-thermophilic digestion process would not be able to handle the entire waste sludge flow without additional capital expenditures for new digestor tanks and separate heating systems, and as such, it was not recommended at this time even though the final product would satisfy all criteria for stabilization and disinfection comparable to effectively operated composting.

4. The amount of grit passing through the existing grit removal facilities is substantial. This grit is combined with the primary sludge and both are pumped to the digestion tanks. The grit accumulates within the digestion tank and reaches equilibrium when about 1/3 of the tank volume is occupied by grit. Accordingly, this grit accumulation has reduced the amount of sludge that can be processed through the existing digestion tanks by at least 1/3.

5. The detailed solids production analysis prepared for this study can be incoprorated into other sludge management evaluations performed by the District of Columbia.

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Acknowledgement

This report has been prepared for the U.S. Environmental Protection Agency, Office of Research and Development, Office of Environmental Engineering Technology (OEET). The work was inspired by the highly successful results obtained under the direction of Mr. Wilbur Torpey at the Rockaway Pollution Control Plant, in New York City. Mr. Torpey served as the principal consultant on this evaluation for the application of the mesophilic-thermophilic process. He also contributed the discussion on the Rockaway test. Mr. James Basilico of OEET, was instrumental in initiating this study and in obtaining the cooperation of the Blue Plains Wastewater Treatment Facility staff in performing the analysis. He was also the USEPA project officer responsible for this study. The detailed analysis of the sludge management system was performed by Environmental Technology Consultants, Inc., under the direction of Mr. Nick Mignone. Dr. John Andrews, University of Houston, assisted in this study as a special consultant on anaerobic digestion, and he also contributed the discussion on the development of the anaerobic digestion processes. WAPORA, Inc., performed the tracer study on the anaerobic digesters and provided overall project management and report preparation under the direction of Mr. Robert France and Mr. Robert Stevens.

Special acknowledgement is due to the District of Columbia management and the Wastewater Treatment Facility staff. In particular to Mr. Steve Bennett and Mr. Ed Jones who provided their time and in-depth understanding of the treatment facility to make the analysis of management alternatives possible, and to Mr. John Zelinski who spent so much time on helping with the process piping.

SECTION 1
INTRODUCTION

This study is an engineering evaluation of the application of a new concept in wastewater engineering to a major wastewater treatment facility. The new concept involves the combination of two anaerobic digestion processes: the mesophilic process (that is, anaerobic digestion operating at temperatures from 90 to 100°F), and the thermophilic process (that is, anaerobic digestion operating at temperatures from 120 to 130°F). In combination, these represent a new process termed the mesophilic-thermophilic process. Such a system has been operated at the Rockaway Pollution Control Plant in New York City with outstanding operational results.

The existing Blue Plains wastewater treatment plant in Washington, D.C. treats approximately 330 mgd and uses mesophilic, anaerobic digestion for sludge treatment. This system has had operational constraints and is limited in capacity. As a result, a major portion of the total sludge stream generated at Blue Plains is disposed of without digestion. Expansions of sludge processing unit operations, now is progress, will increase the system capacity, but not sufficiently to handle the entire sludge stream as the system now operates. Finally, there are related planning and engineering evaluations in progress that consider the long-term sludge disposal options available at Blue Plains, including the complete abandonment of anaerobic digestion in favor of alternate sludge treatment processes.

Therefore, the purpose of these evaluations was to identify the relative merit of the mesophilic-thermophilic process compared with other anaerobic digestion processes. In particular, the evaluations emphasize the capabilities of anaerobic digestion to meet sludge processing needs at Blue Plains, the modifications required to handle the full sludge stream (operating and equipment), and the costs associated with the systems (monetary and energy). Existing equipment is used to the maximum in these evaluations, with the objective to identify an anaerobic sludge digestion process that can be implemented with no major construction. In this sense, the study represents an operations evaluation of the Blue Plains facility for the anaerobic digestion system.

The report is organized to follow logically the analyses performed on the sludge handling system. Section 2 presents a perspective on anaerobic digestion and relevant data and discussions on the new mesophilic-thermophilic process as applied by Mr. Torpey at Rockaway. Section 3 provides a background description of the Blue Plains wastewater treatment facility. Section 4 describes the sludge production at Blue Plains, with detailed information also provided in Appendixes A and B. An operations evaluation of the sludge system is included as Section 5, with supporting analyses in Appendixes C through L. The comparison between anaerobic processes is performed in Section 6, with supporting calculations and assumptions in Appendixes M, N, and O. Financial evaluations are presented in Section 7.

The methods employed in this case study should be generally applicable to most major wastewater treatment facilities. The application of anaerobic processes, and in particular the new mesophilic-anaerobic process, may be profitably analyzed by following these techniques.

SECTION 2
ANAEROBIC DIGESTION AND THE ROCKAWAY STORY

The technologies currently used for the anaerobic digestion of sludge have evolved over the last 100 years as scientific and engineering principles have been applied. Tracing this development, there has been a gradual improvement in performance, both with respect to the stabilization of the waste material and to the associated cost.

2.1. DEVELOPMENT OF THE CONVENTIONAL (MESOPHILIC) PROCESS

Anaerobic digestion is one of the oldest wastewater treatment processes. This process, which involves the biological conversion of organic solids to methane and carbon dioxide, is a natural process occurring in such diverse environments as swamps, stagnant bodies of water, and stomachs of cows. One of its first engineered uses was in the latter part of the 19th century when septic tanks were first used for wastewater treatment. In the septic tank, the solids that settle to the bottom undergo anaerobic decomposition with the liquid passing on to a tile drainage field. The solids are well stabilized, but unfortunately the gas evolved disturbs the sedimentation process by lifting particles into the overflow. This can result in plugging of the tile field, thus destroying the efficiency of the field, and frequently resulting in malodorous conditions.

This deficiency of the septic tank was recognized and a solution developed by the famous sanitary engineer, Dr. Karl Imhoff, in the early part of this century. Dr. Imhoff invented a two-story tank, which now bears his name. The design of this was such that the gas evolved by anaerobic digestion in the bottom tank was prevented from entering the upper tank where sedimentation occurred. The functions of digestion and sedimentation were thus effectively separated.

A natural evolution of this separation of functions, which took place in the 1920's, was the construction of separate tanks for sedimentation and digestion with the solids removed in the sedimentation basin being pumped to the anaerobic digester. This decreased the required depth of the tanks and also permitted the easy application of heating and artificial mixing to the anaerobic digester. Both heating and mixing, which began to be applied in the 1930's and 40's, accelerate the rate at which the solids are digested. They also help to reduce stratification problems in the digester, thus increasing the effective volume available for digestion. This increase in effective volume made possible the use of much smaller digesters. Instead of the three to six months requirement for anaerobic digestion in the Imhoff tank, it was now possible to accelerate digestion and complete the process in one to two months. The obvious result of this was a substantial decrease in required capital cost.

The application of digester mixing soon made it obvious that mixing and separation of the supernatant from the digested sludge were incompatible in much the same sense that sedimentation and digestion in the septic tank were incompatible. This led to a further separation of functions by the application of two-stage digestion where the biological reactions (with mixing and heating for acceleration) occur in the first stage with the digested sludge then going to a separate stage for separation of the solids from the liquid. This second stage also served two other valuable functions, these being the storage of the sludge prior to disposal and as a source of "seed" sludge for restarting the primary digester in the event of difficulties with the biological reactions. Two pioneers in the application of two-stage digestion were A.M. Buswell and A.J. Fisher. With a two-stage system, good digestion could consistently be obtained in the first tank in one month or less.

In the 1950's, Wilbur Torpey was faced with the necessity of expanding the capacity of the digesters used in the New York City plants. He recognized that the digestion time could be substantially decreased if a portion of the water could be removed from the sludge prior to its being fed to the digester. He therefore installed sludge thickeners prior to digestion and was able to quadruple the loading of solids to his digesters. This permitted plant expansion to be put off for several years, thus saving the City of New York a very substantial sum of money. Energy requirements for sludge heating also were substantially reduced since it was no longer necessary to heat the water that was formerly associated with the sludge.

In addition to his full-scale work on thickening, Torpey conducted pilot plant studies that established three days as the lower limit of time at which the process could operate. One would, of course, not normally operate at three days since it allows no margin of safety and digestion is incomplete. Also, the volume of Torpey's pilot digester was fully effective, whereas a portion of the volume of full-scale digesters may be occupied by grit deposits on the bottom or scum at the top. The lower limit of time, with which operating engineers in large cities feel comfortable, is about 15 days.

2.2. PRESENT STATE OF PRACTICE OF THERMOPHILIC ANAEROBIC SLUDGE DIGESTION

Thermophilic anaerobic digestion is very similar to mesophilic anaerobic digestion except for the temperature at which it is operated, this being 120-130° instead of the usual 85-95°F. It thus takes advantage of the well known fact that biological reaction rates can be increased by increasing temperature. It is only natural, therefore, that conversion of existing mesophilic digesters to thermophilic operation should be considered as a low cost technique for increasing the sludge processing capability of wastewater treatment plants. Full scale studies by the Metropolitan Sanitary District of Greater Chicago (1), Ontario Minstry of the Environment, Canada (2), and Moscow, U.S.S.R. (3) all indicated that they could at least double the amount of sludge that could be processed per unit volume of digester capacity by converting from mesophilic to thermophilic operation. The actual amounts of sludge that can be processed may be substantially greater than this since the upper limits of the process have yet to be defined.

In addition to its increased sludge processing capability, thermophilic operation also offers two other significant advantages over mesophilic operation; these being (a) improved sludge dewatering (4,5) and (b) increased destruction of pathogens (3,6).

An example of how sludge dewatering can be improved through the use of thermophilic digestion is afforded by the work of Garber (4) on the vacuum filtration of thermophilic sludge at the Hyperion plant in Los Angeles. He reported a 270% increase in vacuum filter yields with a 48% decrease in coagulant dosage for thermophilic as compared to mesophilic sludge. Improved solids-liquid separation is of substantial value in land application of sludge through decreasing the quantity of wet sludge for disposal and thus lowering the cost of transport to the disposal site.

An example of the increased destruction of pathogens through the use of thermophilic digestion is given in the report of Popova and Bolotina (3) on the practice of thermophilic digestion in Moscow, U.S.S.R. They state "The most essential advantage of this process is the sanitary quality of the thermophilic sludge. According to the sanitary officials of the health department, viable eggs of helminths are absent from such a sludge." This improvement in the sanitary quality of the sludge is of special significance in light of the current trend toward land disposal of digested sewage sludge.

Although mesophilic and thermophilic anaerobic digestion are quite similar in both design and operation, there are differences which must be taken into account in adapting mesophilic digesters to thermophilic operation. Among these are (a) additional sludge heating requirements, (b) structural competency of existing digesters and piping at the higher temperatures, (c) potential odors at sludge handling areas, (d) closer attention to temperature control, (e) possibly higher concentrations of dissolved materials in the liquid streams from sludge dewatering operations, (f) possible ammonia inhibition due to increased protein destruction, and (g) removal of increased amounts of moisture from the digester gas.

It should also be pointed out that caution should be exercised in making the transition from mesophilic to thermophilic operation since the maximum rate at which this can be accomplished is unknown.

PROCESS HISTORY

There has been an interest in thermophilic anaerobic digestion since at least 1930. Buhr and Andrews (7) prepared a review paper of the research aspects and full scale application of the process in 1976 and their paper should be consulted for details of this earlier work. Since 1976, there has been a substantial increase of interest in the process. Different organizations have either adopted or are considering adopting the process to obtain one of the three advantages previously mentioned; these being (a) increased sludge processing rates, (b) improved sludge dewatering, or (c) increased destruction of pathogens. In additon to these three reasons for adopting the process, Metro Denver (8) has converted all of its mesophilic digesters to thermophilic operation in order to compensate for the loss of capacity when the liquid levels were lowered to alleviate problems from the existing foaming conditions. A brief summary of the experiences of some of these organizations is given below.

Los Angeles

The most extensive application of thermophilic anaerobic digestion in the U.S. has been at the Hyperion plant (340 MGD) in Los Angeles. This work has been summarized in a recent paper by Garber (9) with more details of the work being covered in a series of papers by Garber and coworkers (4,5,6,10). The primary advantage of the process for Los Angeles is the greatly improved dewatering characteristics of the sludge.

Thermophilic anaerobic digestion was first started in Los Angeles in 1952 and was subsequently extended to one half of their digestion system. At that time, the digested sludge was vacuum filtered, dried, pelletized, sacked and sold as a soil amendment. The half and half mixture of mesophilic and thermophilic sludge was used since thermophilic sludge has a lower nitrogen content and could not therefore, alone meet the 2.5% nitrogen content guarantee for the soil amendment.

In 1957, Los Angeles commenced discharge of liquid digested sludge to the ocean so dewatering of the sludge was no longer necessary. However, they have continued to operate one or two full sized digesters at thermophilic temperatures so as to gain experience with operational stability and to work on the centrifugability and the pathogen content of the thermophilic sludge. In 1985, Los Angeles will be required to cease discharge of digested sludge to the ocean and turn to land disposal. At that time they plan to operate the entire digestion system at thermophilic temperatures because they will once again need the improved dewatering characteristics.

The work reported on by Garber and coworkers in 1975 (5) and 1977 (6,10) confirmed their earlier work in 1954 in illustrating that solids dewatering (both vacuum filtration and centrifugation) was definitely benefited by thermophilic operation. They also found that the process was no more difficult to operate than the mesophilic process except that closer temperature control was required. In a cooperative study with the Municipal Environmental Laboratory of the Environmental Protection Agency (11), they also found that thermophilic operation resulted in very substantial improvements in the destruction of pathogenic bacteria and viruses. For example, thermophilic digestion consistently reduced the Salmonella densities to below the detectable limits of the analytical procedure.

The volatile acids concentration in thermophilic digesters is higher than that in mesophilic digesters and this was also the case in Los Angeles. The odor of thermophilic sludge is accordingly also somewhat higher. Although Garber did not feel that this was at a level which would be considered obnoxious, he did suggest that care should be taken in the solids dewatering step to contain these odors. The Los Angeles experiences also indicate that there are somewhat higher concentrations of dissolved materials in the liquid streams from dewatering operations.

Moscow, U.S.S.R.

In 1964, Popova and Bolotina (3) reported on the use of thermophilic digestion at the Kur'yanova plant (260 MGD) in Moscow. Plant scale tests on thermophilic operation were first made at this plant in 1955 and in 1958 the digestion system was converted to thermophilic operation. The primary reason for this conversion was the sanitary quality of the thermophilic sludge. Research at the plant had shown that mesophilic digested sludge retained up to 20% viable helminth eggs whereas after thermophilic digestion no viable eggs could be found. The sanitary quality of the sludge was of particular concern since it had become the practice to apply the liquid sludge directly to agricultural land.

Conversion from mesophilic to thermophilic digestion also permitted a sharp increase in the capacity of the installation with the digestion time being shortened from 18 to 9 days.

Chicago

In 1977, the Metropolitan Sanitary District (MSD) of Greater Chicago converted one of their 12 digesters at the West Southwest plant (1,200 MGD) to thermophilic operation. The reason for this is that their mesophilic digesters are currently operating at a 14 day detention time and they wished to explore the possibility of increasing sludge processing capacity by operation in the thermophilic range. Although their experimental program is not completed, they did report on the results of 130 weeks of operation at the 1980 Conference of the Water Pollution Control Federation in Las Vegas (1).

Their work did demonstrate that thermophilic digesters could be operated successfully at a 7 day detention time. Of particular concern to MSD is the energy self-sufficiency of the process since thermophilic digestion does require more heat energy input than the mesophilic process. However, they did observe increased gas production from the thermophilic process and believe that this, with the use of a heat exhanger to preheat the feed sludge with the discharged thermophilic sludge, will offset the increased energy requirements. Just as in Los Angeles, they had no particular difficulty in operating the process and observed that it could stand a change in temperature of up to 5°F in 24 hours without adverse effects.

MSD is particularly concerned with sludge odor since the digested sludge is initially lagooned at a location with a high visability. Their tests indicated only a slightly greater odor intensity from the thermophilic sludge, however, the odor was different and there were various opinions as to whether the ~~order~~ odor was better or worse. They therefore plan further investigations of the disposal aspects of the thermophilic sludge.

Ontario Ministry of the Environment, Canada

In 1977, Smart and Boyko (2) of the Ontario Ministry of the Environment, Canada, reported on the results of their full scale studies on the thermophilic anaerobic digestion process. They had originally planned to establish the upper limits at which sludge could be processed by thermophilic digestion but due to difficulties with their heating system were unable to do so. However, they did obtain good overall performance at a detention time of 7.5 days and stated that the thermophilic digestion process could be applied in existing plant operations where solids stabilization performance is compromised by organic or hydraulic overloading conditons.

As a practical matter, they found that standard gas drying equipment was inadequate for drying the gas produced in the thermophilic digester since it is at a higher temperature and thus contains more water vapor. Several modifications of this equipment were recommended for accomplishing additional moisture removal.

2.3. DEVELOPMENT OF THE MESOPHILIC-THERMOPHILIC PROCESS

Torpey, in his most recent (1979-80) work at the Rockaway plant in New York City, has found a way to overcome the potential disadvantages of thermophilic digestion as well as to improve upon the process. He has accomplished this by again going to two-stage digestion but this time with the two stages consisting of a mesophilic stage followed by a thermophilic stage. He also recycles a portion of the thermophilically digested sludge back through the aeration tanks to obtain additional destruction of organic solids. The advantages of the thermophilic process are thus retained with the potential disadvantages being decreased. In addition, a substantial improvement in organic solids destruction is obtained thus improving the quality of the ultimate product, the digested sludge. The extra gas produced from the destruction of these solids also helps to offset the higher heating requirements. However, a disadvantage of the process is that it does require greater digester capacity (as compared to thermophilic digestion).

2.4 PURPOSE AND METHODS OF THE ROCKAWAY TESTS

The City of New York, which is under Federal mandate to cease ocean dumping of sludge derived from the treatment of wastewater, has been involved in studies of land-based disposal techniques for the past few years. These studies were initially directed towards determination of the optimal methods for dewatering digested sludge and identification of the subsequent steps for ultimate land disposal.

Later the idea was advanced that, as a fundamental and top priority item in the management program, current plant facilities should be studied for use in the reduction of the rate of sludge production, both as to volume and as to volatile matter, from an activated sludge plant. The method would be based on the exploitation of biochemical mechanisms, namely by exposing the mesophilically digested solids to the enzyme systems of thermophilic digestion and activated sludge. A new method of sludge processing has been formulated to implement this idea. The advantage gained in this investigation would be reflected commensurately in the economics of all sludge dewatering and post-dewatering processes which were previously studied.

The Rockaway Pollution Control Plant (P.C.P.), which is connected to a population of 100,000 and has a flow of 25 to 30 mgd, was chosen for a full scale test. The plant uses conventional facilities for the activated sludge process and the sludge generated undergoes mixed primary and seconary sludge thickening prior to mesophilic digestion. The digested sludge is transported to sea. As presently operated, the primary tanks provide a detention of about 2 hours, the aeration tanks of 3.3 hours with step feed provisions, and the final tanks of 3 to 5 hours depending on the number in use. Two 45 ft. diameter thickening tanks are used for mixed sludge thickening and a 1 cu.ft./capita tank is used for the mesophilic digestion.

As required for this test, the following steps were taken: (1) an additional 1 cu.ft./capita digestion tank was placed in service to receive the overflow sludge from the mesophilic digestor and its contents were heated to 120 - 122°F, the lower limit of the thermophilic digestion range; (2) piping was installed to carry a portion of the overflow from the thermophilic digestor directly to a single 45 ft. diameter tank, which was to be utilized as a rethickening and elutriating tank; (3) piping was installed to carry the remainder of the flow from the thermophilic digester into the primary effluent, and thereby, directly into the aerator of the secondary treatment system; (4) city water was carried to the elutriation tank. In summary, the elements of this new method of sludge processing involved (1) subjecting the mesophilically digested sludge to subsequent thermophilic digestion, (2) circulating a portion of the sludge leaving the thermophilic digester directly to, and through, the secondary treatment system, and (3) subjecting the remainder of the sludge leaving the thermophilic digester to a rethickening and elutriation step.

The thermophilic digester was placed in operation in September 1979. However, it was not until January 15, 1980 that the necessary piping additions were completed and the recirculaation and rethickening steps of the new method brought into service. Therefore, although the thermophilic digester was in operation, test results on treatment plant effectiveness before January 1980 do not include any effects due to recirculation.

2.5. RESULTS OF THE ROCKAWAY TESTS

2.5.1 Effect of Recirculation on Process Performance (Solids and BOD)

When the full scale test was started, the activated sludge was at a low sludge density index of 0.6 to 0.7. Microscopic examination revealed a significant population of bacterial filaments present along with colonies of stalk ciliates and rotifers. After only a few days of operation, the digested sludge recirculation caused the sludge density index to rise to

1.0 and the bacterial filaments to diminish substantially. Throughout the entirety of the test following, the sludge density index remained in the stable range of 1.0 to 1.4.

Since the flow received at the plant is approximately 200 gals/capita/day, the influent wastewater strength is low, averaging about 100 mg/l each of suspended solids and BOD_5. Prior to the test, the suspended solids and BOD_5 in the effluent averaged 14 and 12 mg/l respectively. The monthly treatment results for the period from July to December 1979 are presented in Table 2-1. For comparative purposes, the treatment effectiveness during the course of the test from January 15 through May 29, 1980, is also included. These data indicate that no significant effect on treatment efficiency was experienced as a result of the continuous recirculation of digested sludge through the aeration system for at least the first three months. In the latter 2 months, suspended solids in the effluent did increase by about 3 mg/l and there was a substantial increase in the flow rate from about 20 mgd to 27-29 mgd.

2.5.2. Nutrient Removal

To further evaluate whether the recirculation of digested sludge through the secondary system had an adverse effect on effluent quality, the data concerning nitrogen and phosphorous are shown in Tables 2-2 and 2-3. The effluent values are of special interest since the raw wastewater samples did not contain the recirculating flow. Inspection of the data for the two periods (1) pre-test and (2) test, shows that the total average effluent inorganic nitrogen was 9.2 mg/l and 10.4 mg/l respectively. Although data from month to month varies, period averages show that inorganic nitrogen in the effluent increased 1.2 mg/l during the test period over the pre-test period. Conversely, the organic nitrogen decreased 4.3 mg/l. Phosphorus concentrations in the effluent remained essentially unaffected over the two periods. It appears that the digested sludge recirculation had a minor effect on the effluent quality with respect to the nutrients nitrogen and phosphorus.

TABLE 2-1. ROCKAWAY P.C.P. TREATMENT EFFICIENCY, JULY 1979 TO MAY 1980

		Plant Influent				Plant Effluent			
	Flow	Suspended Solids		BOD_5		Suspended Solids		BOD_5	
Month	MGD	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day
Before Modification									
July 79	22	88	16,146	111	20,366	12	2202	13	2385
Aug	22	83	15,289	117	21,467	16	2936	12	2202
Sept	23	93	17,839	90	17,264	12	2302	10	1918
Oct	25	116	24,186	103	21,475	18	3753	12	2502
Nov	22	140	25,687	112	20,550	13	2385	11	2018
Dec	21	125	21,892	124	21,717	10	1751	11	1927
AVG		108	20,173	110	20,473	14	2555	12	2159
After Modification									
Jan 80	21	86	15,062	91	15,938	8	1401	8	1401
Feb	19	86	13,628	85	13,469	9	1426	8	1268
Mar	23	106	20,333	57	10,934	13	2494	8	1534
Apr	27	94	21,167	49	11,034	15	3378	6	1351
May	29	94	22,734	43	10,400	15	3628	6	1451
AVG		94	18,585	65	12,355	12	2465	7	1401

TABLE 2-2. ROCKAWAY P.C P. INFLUENT-EFFLUENT NITROGEN CONCENTRATIONS, JULY 1979 TO MAY 1980

Before Modification

Month	Flow MGD	Ammonia Nitrogen Influent mg/l	Ammonia Nitrogen Influent lbs/day	Ammonia Nitrogen Effluent mg/l	Ammonia Nitrogen Effluent lbs/day	Organic Nitrogen Influent mg/l	Organic Nitrogen Influent lbs/day	Organic Nitrogen Effluent mg/l	Organic Nitrogen Effluent lbs/day	Nitrate Nitrogen Influent mg/l	Nitrate Nitrogen Influent lbs/day	Nitrate Nitrogen Effluent mg/l	Nitrate Nitrogen Effluent lbs/day	Total Inorganic Nitrogen Influent mg/l	Total Inorganic Nitrogen Influent lbs/day	Total Inorganic Nitrogen Effluent mg/l	Total Inorganic Nitrogen Effluent lbs/day
1979																	
July	22	13.0	2385	1.6	294	5.8	1064	2.9	532	0	0	7.0	1284	13.0	2385	10.6	1945
Aug	22	7.8	1431	1.8	330	8.4	1541	3.0	550	2.3	422	5.9	1082	10.1	1853	9.7	1180
Sept	23	----[a]	----	---	----	10.5	2014	9.6	1841	---	---	---	----	----	----	----	----
Oct	24	7.8	1626	0.6	125	8.4	1751	3.0	625	0	0	5.2	1084	7.8	1626	6.0	1251
Nov	22	12.5	2294	9.4	1724	15.1	2771	8.6	1578	0.1	18	0.4	73	12.8	2348	10.3	1890
Dec	21	10.4	1821	9.2	1611	16.0	2802	14.8	2592	0.3	52	0.2	35	10.7	1874	9.4	1646
AVG		10.3	1911	4.5	817	10.7	1991	7.0	1050	---[b]	--[b]	3.7	----[b]	10.9	2017	9.2	1582

After Modification

Month	Flow MGD	Ammonia Nitrogen Influent mg/l	Ammonia Nitrogen Influent lbs/day	Ammonia Nitrogen Effluent mg/l	Ammonia Nitrogen Effluent lbs/day	Organic Nitrogen Influent mg/l	Organic Nitrogen Influent lbs/day	Organic Nitrogen Effluent mg/l	Organic Nitrogen Effluent lbs/day	Nitrate Nitrogen Influent mg/l	Nitrate Nitrogen Influent lbs/day	Nitrate Nitrogen Effluent mg/l	Nitrate Nitrogen Effluent lbs/day	Total Inorganic Nitrogen Influent mg/l	Total Inorganic Nitrogen Influent lbs/day	Total Inorganic Nitrogen Effluent mg/l	Total Inorganic Nitrogen Effluent lbs/day
1980																	
Jan	21	9.4	1646	0.6[c]	105[c]	9.8	1716	3.0[c]	525[c]	0.2	35	6.3[c]	1103[c]	9.6	1681	6.9[c]	1208[c]
Feb	19	11.6	1838	2.6	412	10.6	1680	2.2	349	0.3	48	8.2	1299	11.9	1886	10.8	1711
Mar	23	9.6	1841	3.0	575	15.6	2992	4.2	806	0.8	153	10.0	1918	10.4	1995	13.2	2532
Apr	27	7.0	1576	1.0	225	8.6	1937	3.2	721	0.3	68	5.6	1261	7.3	1645	9.8	2207
May	29	9.4	2273	2.4	580	9.2	2225	1.2	290	0.3	73	5.4	1306	9.7	2346	7.8	1887
AVG		9.4	1835	2.3	448	10.8	2110	2.7	542	0.4	75	7.3	1446	9.8	1911	10.4	2084

[a] Analytical results not available.

[b] Averages not calculated

[c] Transition month-results not included in averages.

TABLE 2-3. ROCKAWAY P.C.P. INFLUENT-EFFLUENT PHOSPHORUS CONCENTRATIONS, JULY 1979 TO MAY 1980

		Total Phosphorus				Ortho Phosphorus			
	Flow	Influent		Effluent		Influent		Effluent	
Month	MGD	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day
Before Modification									
July 79	22	2.3	422	1.8	330	1.7	312	1.8	330
Aug	22	2.5	459	2.1	385	1.8	330	1.4	257
Sept	23	2.5	480	2.1	403	1.1	211	1.7	326
Oct	25	2.0	417	1.2	250	1.4	292	1.2	250
Nov	22	2.7	495	1.6	293	1.9	349	1.2	220
Dec	21	1.8	315	1.6	280	3.5	613	3.0	525
AVG		2.3	431	1.7	323	1.9	351	1.7	318
After Modification									
Jan 80	21	2.7	473	1.8[a]	315[a]	1.6	280	1.5[a]	262[a]
Feb	19	2.8	444	1.7	269	1.8	285	1.4	222
Mar	23	3.9	748	2.0	384	1.9	364	0.7	134
Apr	27	2.9	653	2.4	540	1.2	270	1.1	248
May	29	2.5	604	2.0	484	1.3	314	1.6	387
AVG		3.0	584	2.0	419	1.6	303	1.2	248

[a] Transition month results not included in averages.

2.5.3. Heavy Metals Removal

The results of the monthly metal analyses of composite influent and effluent samples for the pre-test period July to December 1979 and for the test period January to May 1980 are presented in Table 2-4. Comparison of the overall averages of these test periods for cadmium and mercury, the two metals that have been shown to exert a major effect on human physiology, indicate that there is not a significant difference between the removals. In fact, the activated sludge process does not seem capable of appreciably reducing the already low concentration of either of these metals. Also, mass balance studies of the metal data, excepting cadimum and mercury, have been generally good. Cadmium and mercury, however, due to their low concentrations and to the sensitivity of the testing procedure, had poor mass balances. Comparative inspection of the data for the remaining heavy metals, indicates some variable effects of treatment during individual months, but with no significant changes in the overall averages for the subject periods.

2.5.4. Effect of Recirculation on Oxygen Requirements

The effect of digested sludge recirculation on air compressor output over the course of the test could not be determined. The air compressor was operating at a level to produce more than adequate dissolved oxygen and its rate could not be lowered before or during the test to attempt to evaluate any demand changes. A calculated estimate was made based on the fact that meso-digestion, without the benefit of subsequent thermo-digestion, destroys 90% of the BOD_5 present. The estimate indicates that the BOD_5 in the portion of the digested sludge which is continuously recirculated, would add less than 5% to the oxygen demand of the primary effluent.

TABLE 2-4. ROCKAWAY P C.P. INFLUENT-EFFLUENT METALS CONCENTRATIONS, JULY 1979 TO MAY 1980

	Copper		Chrome		Nickel		Zinc		Lead	
	Inf	Eff	Inf	Eff	Inf	Eff	Inf	Eff	Inf	Eff
Before Modification										
July 1979	0.11	0.07	0.001	0.006	0.07	0.04	0.11	0.14	0.017	0.006
Aug	0.11	0.05	0.020	0.015	0.02	0.02	0.26	0.35	0.024	0.006
Sept	0.13	--	0.012	0.008	0.02	0.01	0.23	0.16	0.110	0.006
Oct	0.13	0.03	0.009	0.002	0.01	0.01	0.09	0.17	0.014	0.008
Nov	0.16	0.06	0.011	--	0.02	0.02	0.12	0.15	0.027	0.020
Dec	0.11	0.05	0.034	0.007	0.01	0.02	0.08	0.08	0.023	0.007
AVG	0.12	0.05	0.014	0.008	0.03	0.02	0.15	0.17	0.036	0.009
After Modification										
Jan 1980	0.095	0.22	0.007	0.0012	0.015	0.018	0.066	0.070	0.014	0.030
Feb	0.080	0.0035	0.0012	0.001	0.009	0.021	0.093	0.086	0.049	0.0024
Mar	0.110	0.0400	0.0038	0.003	0.0042	0.0024	0.090	0.065	0.0089	0.0016
Apr	--	--	0.010	0.005	0.0068	0.014	0.21	0.079	0.0088	0.0034
May	0.075	0.0380	0.0038	0.001	0.0086	0.011	0.085	0.10	0.010	0.0064
AVG	0.088	0.027	0.0047	0.002	0.0072	0.012	0.12	0.082	0.018	0.0035

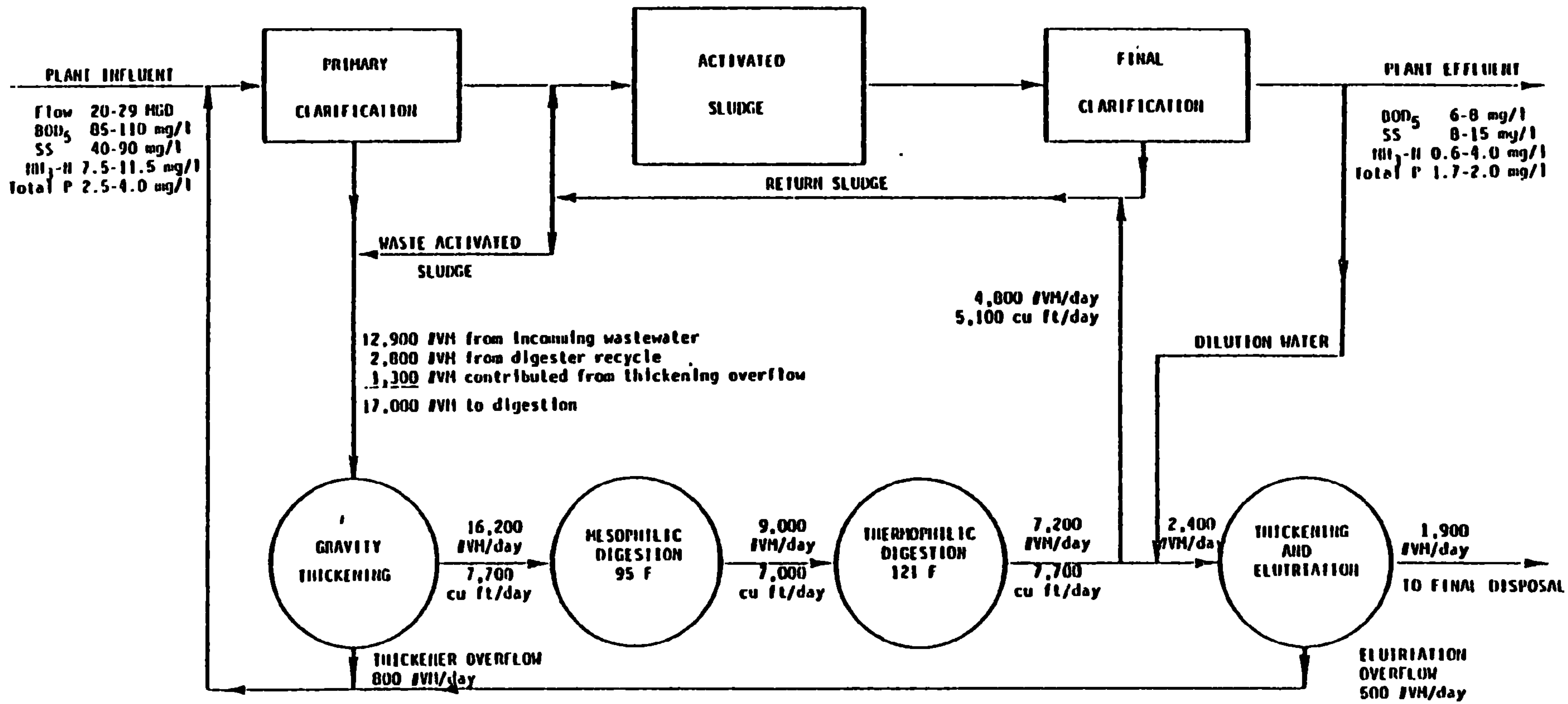
	Iron		Cadmium		Calcium		Magnesium		Mercury	
	Inf	Eff	Inf	Eff	Inf	Eff	Inf	Eff	Inf	Eff
Before Modification										
July 1979	0.8	0.2	0.0001	0.0001	41	36	94	98	0.0010	0.0006
Aug	0.9	1.0	0.0001	0.0001	40	48	107	112	0.0007	0.0006
Sept	1.5	0.7	0.004	0.0002	41	32	102	105	0.0007	0.0002
Oct	1.5	1.2	0.0001	0.0001	29	41	94	101	0.0005	0.0005
Nov	1.6	2.0	0.0010	0.0015	19	20	85	88	0.0026	0.0028
Dec	1.1	0.1	0.0008	0.0006	13	15	76	77	0.0005	0.0009
AVG	1.2	0.9	0.0004	0.0004	30	32	93	97	0.0010	0.0009
After Modification										
Jan 1980	0.57	1.00	0.0018	0.0005	14	14	70	72	0.0009	0.0011
Feb	0.65	0.16	0.0005	0.0004	16	17	60	64	0.0005	0.0003
Mar	0.55	0.21	0.0011	0.0009	14	15	58	59	0.0009	0.0002
Apr	0.84	0.13	0.0046	0.0029	13	14	54	55	0.0003	0.0005
May	0.83	0.46	0.0011	0.0017	23	21	59	62	0.0003	0.0004
AVG	0.72	0.24	0.0018	0.0015	16	17	58	60	0.0005	0.0004

2.5.5. Operating Results

As was previously noted, a 1 cu. ft./capita tank was placed in service as a thermo-digester at about 121°F. Its contents overflowed by gravity to either the primary effluent, or to a 45 ft. diameter rethickening and elutriating tank where city water was added to the influent sludge at a ratio of about 3 to 1. The rethickened underflow sludge was pumped by a duplex plunger pump, actuated by time clock, to empty digesters where its volumetric rate was measured by filling the tanks during the months of March and April. Such procedures did not involve the use of any man power, except for periodical blowing back of clogged lines.

The amount and concentration data are presented in Table 2-5. The volatile matter leaving the meso-digester averaged 9,000 lbs/day, thus effecting a reduction of 7,200 lbs/day (16,200 - 9,000). The thermo-digester accounted for a further reduction of 1,800 lbs/day (9,000 - 7,200). Such reductions were effected on the combination of raw primary solids, activated sludge solids and the recirculating solids (i.e., solids that had been previously subjected to meso and thermo-digestion).

The conventional activated sludge treatment units and the modifications made for incorporating the new method of reducing sludge production are represented in Figure 2-1. Since the economics of sludge disposal is fundamentally a function of the amount of volatile material to be disposed of, only the rates of production of volatile matter are shown. Overall, it can be seen that the reduction of volatile matter by the meso-digester of 7,200 lbs/day, added to the 1,800 lbs/day reduction by the thermo-digester, results in a total of 9,000 lbs/day. Additionally, the aerator destroyed 2,000 lbs. V.M./day for a total reduction of 11,000 lbs.V.M./day. Since the treatment system was removing a total of 12,900 lbs.V.M./day, the net amount requiring disposal was reduced to 1,900 lbs.V.M./day. Previous data show the average amount of volatile matter carried to sea from April to July 1979 (the period just prior to this work) was 5,700 pounds. Therefore, there was a reduction of two-thirds in the volatile solids requiring disposal ($\frac{5700 - 1900}{5700} = 2/3$). Volume reduction was in the same proportion; that is, 4,800 cu ft/day to 1,650 cu ft/day, or approximately 2/3.



VOLATILE MATTER TO DISPOSAL

Before thermophilic system incorporated	After thermophilic system incorporated
5,700 #, 4,800 cu ft	1,900 #, 1,650 cu ft

#VM/day - POUNDS VOLATILE MATERIAL PER DAY

FIGURE 2-1. Summary of Operating Results

TABLE 2-5. ROCKAWAY P.C.P. AMOUNT AND CONCENTRATION OF SOLIDS PASSING THROUGH SYSTEM

				%Conc. V.M.				#V.M./day leaving			
										Rethickener Elutriator	
Mo. 1980	Flow MGD	#VSS capt. (75% VM)	Raw thick. pump. cu ft/day	Raw thick	Meso dig.	Thermo dig.	#V.M. from thick.	Meso dig.	Thermo dig	Under flow	Over flow
Jan	21	10,400	5,900	3.9	1.6	1.1	14,500	6,000	4,100	-	-
Feb	19	9,500	7,300	3.6	1.8	1.3	16,500	8,200	5,900	-	-
Mar	23	14,800	8,400	3.3	1.7	1.5	17,500	8,900	7,800	1,800	500
Apr	27	13,400	6,500	3.5	1.9	1.5	14,500	8,000	6,400	2,000	400
May	29	14,200	8,500	3.0	2.0	1.6	16,200	10,800	8,600	-	600
Avg. Feb to May	25	12,900	7,700	3.4	1.9	1.5	16,200	9,000	7,200	1,900	500

Note: (1) Calc. of V.M. inventory in digesters after February show the change does not significantly influence the above data.

(2) Measured volume March 1,600 cu ft/day x 63 x 1.8% = 1,800 #V.M./day
April 1,700 cu ft/day x 63 x 1.9% = 2,000 #V.M./day

The daily amount of gas generated during the course of the thermo-digestion is shown in Table 2-6. Based on the averages for the period February to May, the meso-digester produced 83,900 cu ft/day gas, slightly less than the comparable preceding period without digested sludge recirculation through the aerator. The gas generated by the thermo-digester increased from an average of 7,000 cu ft/day to 14,000 cu ft/day during recirculation. Gas production per pound of volatile solids reduced is presented in Table 2-7, with gas production for several pure compounds presented in Table 2-8 for comparison.

The gas mixers in both digesters were causing formation of large solids masses in the digesters with consequent clogging of the overflows. It was necessary to operate the mixers only a few minutes per day to alleviate this condition.

Garber (1) in Los Angeles had determined that the thermo-digested sludge required half the dose of iron coagulant and produced almost four times the yield on a vacumm filter as meso-digested sludge. Accordingly, to obtain some estimate of the improvement on coagulability achieved by the use of the thermo-digestion in this instance, the meso and thermo-digested sludges were subjected to polymer treatment. It was found on a laboratory scale, that using a high molecular weight, low charge polymer #2535CH (as manufactured by American Cynamid) the coagulability improved radically. Specifically, dosages of up to 4,000 ppm on meso-digested sludge did not produce an end point, although some flocculation was observed. In contrast, the thermo-digested sludge released 73% of the water in 30 minutes at 1G, at a dose of 2,500 ppm. Thermo-digested sludge, after a 3 to 1 elutriation, required a lesser comparative dose of 1,650 ppm of the same polymer to release 64% of the water within 30 minutes at 1G.

TABLE 2-6. ROCKAWAY P.C.P. DAILY GAS PRODUCTION

Month	Mesophilic digester cu ft/day	Thermophilic digester cu ft/day
September 1979	79,200	5,300
October	93,800	8,200
November	90,300	6,900
December	77,200	7,500
Average, No recirculation	87,600	7,000
January 1980	79,200	8,500
February	88,000	12,700
March	86,800	13,600
April	76,400	11,900
May	84,300	17,700
Average, with recirculation Feb to May 1980	83,900	14,000

TABLE 2-7. GAS PRODUCTION CUBIC FEET/POUND VOLATILE SOLIDS REDUCED

Month	Mesophilic digester	Thermophilic digester
January	9.3	4.5
February	10.6	5.5
March	10.1	12.4
April	11.7	7.4
May	15.6	8.4

TABLE 2-8. GAS PRODUCTION AND QUALITY FROM ANAEROBIC DIGESTION OF SEVERAL PURE COMPOUNDS

Material	Gas production cubic feet/pound volatile solids reduced	Percent content
Crude fibers	13	45-50
Fats	18-23	62-72
Grease	17	68
Protein	12	73
Scum	14-16	70-75

SUMMARY

A full scale test was conducted at the Rockaway Plant in New York City, which serves a population of 100,000, for a period of five months to evaluate a new method of reducing the amount and volume of sludge produced from the activated sludge process. This method involved the use of (1) high stability thermophilic digestion following mesophilic digestion, and (2) the recirculation of a portion of such thermo-digested sludge directly to and through the secondary system of the activated sludge process while the remainder was conducted to a rethickening and elutriation tank. Operating results have demonstrated that the volatile matter normally transported to sea after meso-digestion was reduced by 2/3. Moreover, the volume of sludge produced was lowered by 2/3 without chemical or mechanical aids. Using a laboratory scale, it was shown that the residual solids exhibited improved coagulability after having undergone thermo-digestion. This change would improve the economics of all subsequent dewatering processes. The treatment process was performed without significant adverse effect on any accepted parameter due to the continuing recirculation of digested sludge through the activated sludge process.

Reference: (1) Buhr, H. O. and Andrews, J. - The Thermophilic Anaerobic Digestion Process - Water Research. Vol. II, pp139-146, 1977, Permanon Press, Great Britain.

SECTION 3

BLUE PLAINS WASTEWATER TREATMENT FACILITY

The District of Columbia WWTF is located in the southernmost portion of Washington, DC, on the east bank of the Potomac River. The treatment facility receives flow from the District of Columbia, Virginia, and Maryland. The plant has a tributary area of 725 square miles, with a design flow rate of 309 mgd and a peak flow rate of 650 mgd. It occupies an area of more than 200 acres, and is adjoined by the US Naval Research Laboratory to the north, the Anacostia Freeway to the east, and the Potomac River to the south and west.

The existing (winter 1980) liquid management treatment scheme consists of raw wastewater pumping, aerated grit removal, primary clarification, high rate activated sludge, intermediate clarification with chemical addition for phosphorus removal, nitrification, final clarification, multimedia filtration, and chlorine disinfection. Discharge is to the Potomac River.

The existing (winter 1980) sludge management treatment scheme consists of the following. All primary sludge is thickened in gravity thickeners; all waste activated, waste nitrified, and filter backwash sludge is thickened in dissolved air flotation thickeners. Approximately half of the total sludge volume is pumped to the existing anaerobic digesters, elutriated, vacuum filtered, and disposed of on agricultural land. The other half of the total sludge volume is vacuum filtered with excess lime and is either landfilled in trenches or composted.

The size and capacities of the existing sludge handling equipment are discussed in Section 5.

SECTION 4
EVALUATION OF CURRENT AND FUTURE PRIMARY AND SECONDARY SLUDGE GENERATION

The first step in this analysis of ways to improve the existing anaerobic digestion operation at Blue Plains involved the evaluation of current and future sludge production. Current primary and secondary sludge production was determined using operation data supplied by Blue Plains personnel for the months of January through June 1980 (Appendix A, Table A-1). Future primary and secondary sludge production was estimated by using existing data plus assumptions as to the effects on sludge production of the completion of current on-going plant modifications. All data analyses relative to sludge production are given in Appendix B.

4.1. PRIMARY SLUDGE PRODUCTION

Primary sludge consists of those solids that settle and are removed in the primary clarifiers. Currently, these solids are generated from the raw plant influent after coarse screening and grit removal and the overflow from the gravity thickening and elutriation tanks. Analysis of the data indicates that the average dry solids withdrawn from the primary clarifiers and pumped to the gravity thickeners is 0.348 ± 0.057 dry tons of solids per million gallons of influent flow (T/MGIF).[1] The volatile solids content averages 79.7 ± 2.9 percent or 0.277 ± 0.047 dry tons of volatile solids per million gallons of influent flow (VT/MGIF).

Primary sludge production and volatile content are not expected to change by any significant amount within the near future.

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[1] It is estimated that 0.07 to 0.1 T/MGIF is grit, which is not being removed in the existing aerated grit chambers.

4.2. SECONDARY SLUDGE PRODUCTION

Secondary sludge currently consists of those solids that settle out in the intermediate clarifiers and are removed from the system as waste sludge. Currently, these solids come from the biological solids generated in the high rate, activated sludge system; the suspended solids that did not settle out in the primary clarifiers; and the solids produced from the addition of iron for phosphorus removal. Analysis of the data indicates that the average dry solids wasted from the intermediate clarifiers and pumped to the dissolved air flotation thickeners is 0.284 ± 0.034 T/MGIF.[2] The volatile solids average 65.3 ± 2.1 percent or 0.185 ± 0.022 VT/MGIF.

Secondary sludge production will increase in the near future for three reasons: increased phosphorus removal, operation of nitrification systems, and operation of multi-media filters.

4.2.1. Increased Phosphorus Removal

Effluent limitations on phosphorus will need to be reduced from the current 1.1 mg/l average to the NPDES permit requirement of 0.53 mg/l. This will increase chemical sludge production from iron addition by approximately 0.009 T/MGIF, assuming that the volatile fraction is negligible.

4.2.2. Nitrification

At a minimum, this system will have to reduce current total Kjeldahl nitrogen (TKN) levels in the plant effluent from an average of 13.8 mg/l to 5.3 mg/l. Assuming a biological yield coefficient of 0.1 pounds of solids produced per pound of TKN reduced, nitrification would generate an additional 0.004 T/MGIF, of which 70 percent is assumed to be volatile.

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[2] Approximately 0.045 ± 0.013 T/MGIF is attributed to chemical sludge generated by iron addition for phosphorus removal.

4.2.3. Multi-media Filters

By the end of 1981, the new multi-media polishing filters will be in operation. It is estimated that these filters will reduce the suspended solids being discharged from the current 15 mg/l average to 7.5 mg/l. This will contribute approximately 0.031 T/MGIF of which 70 percent is assumed to be volatile.

These three new sources will increase secondary sludge production to approximately 0.328 ± 0.034 tons of dry solids per MG of raw influent flow. The volatile solids content of the secondary sludge would be reduced to 0.210 ± 0.023 VT/MGIF or 63.8 ± 2.1 percent.

4.3. SUMMARY

Table 4-1 summarizes the information presented in Sections 4.1 and 4.2. Future sludge production should be realized by the fall of 1981.

TABLE 4-1. SUMMARY OF CURRENT AND FUTURE PRIMARY AND SECONDARY SLUDGE GENERATION AT BLUE PLAINS

	Current sludge generation		Future sludge generation	
Sludge type	Dry tons (pounds) per million gallons of influent flow	Volatile dry tons (pounds) per million gallons of influent flow	Dry tons (pounds) per million gallons of influent flow	Volatile dry tons (pounds) per million gallons of influent flow
Primary	0.348 ± 0.057 (696 ± 114)	0.277 ± 0.047 (554 ± 94)	0.348 ± 0.057 (696 ± 114)	0.277 ± 0.047 (544 ± 94)
Secondary				
o High rate system	0.239 ± 0.034 (478 ± 68)	0.185 ± 0.022 (370 ± 44)	0.239 ± 0.034 (478 ± 68)	0.185 ± 0.022 (370 ± 44)
o Chemical	0.045 ± 0.013 (90 ± 26)	------------------------	0.054 ± 0.013 (108 ± 26)	------------------------
o Multi-media	------------------------	------------------------	0.031 (62)	0.022 (44)
o Nitrification	------------------------	------------------------	0.004 (8)	0.003 (6)
TOTAL	0.632 ± 0.068 (1264 ± 136)	0.462 ± 0.051 (924 ± 102)	0.676 ± 0.068 (1352 ± 136)	0.487 ± 0.051 (974 ± 102)

SECTION 5

EVALUATION OF EXISTING SLUDGE MANAGEMENT OPERATIONS ON SLUDGE PROCESSING AND SLUDGE GENERATION

The second step in the analysis of ways to improve the existing anaerobic digestion operation at the District of Columbia WWTF was to evaluate existing sludge management operations. This section evaluates operating data supplied by Blue Plains personnel for the months of January through June 1980 (Appendix A). Data analysis, where required, is provided in the appendix specified in the following discussion.

5.1. GRAVITY THICKENER OPERATION

There are six, 65-foot diameter gravity thickeners. Either primary or secondary sludge can be pumped to these units, but since the installation of the dissolved air flotation thickeners, only primary sludge has been thickened in these tanks. Thickened sludge and/or scum collected from the tank surface can be pumped either to the anaerobic digestion system or directly to raw sludge dewatering. Overflow from the thickeners is returned to primary treatment. Operational data for the months of January through June 1980 is provided in Appendix A, Table A-2. Data analysis is provided in Appendix C.

Current practice is to pump primary sludge 24 hours per day from the primary clarifiers to the gravity thickeners at a solids concentration of less than one percent. In addition, part of the elutriation tank overflow is continually returned to the gravity thickeners. Dilution water from the intermediate clarifier overflow is added. Experience has indicated that the rate of dilution water utilization should be 800 gallons per day per square foot of thickener tank surface. Data analysis indicates that under this mode of operation:

1. The average thickened solids concentration that can be expected is 7.0 ± 0.5 percent.

2. The average thickened sludge volume per million gallons of plant influent flow is 1,236 ± 218 gallons.

3. The amount of and volatile content of primary sludge produced per million gallons of influent flow is altered. After gravity thickening, the average dry solids withdrawn is 0.377 ± 0.060 dry tons of solids per million gallons of influent flow (T/MGIF).[1] The volatile solids content averages 71.0 ± 5.3 percent or 0.268 ± 0.047 dry tons of volatile solids per million gallons of influent flow (VT/MGIF).

Future performance of the gravity thickeners could change significantly, depending on the following conditions:

1. Change in the amount of grit that is currently getting into the primary sludge stream.

2. Change in the source and amount of dilution water currently being used.

3. Change in the quantity of elutriation tank overflow presently being sent to the gravity thickeners.

4. Change in the type of sludge thickened i.e. thickening a mixture of waste primary and secondary sludge solids.

5.2. DISSOLVED AIR FLOTATION (DAF) THICKENER OPERATION

There are eighteen, 20-foot wide by 55-foot long (effective length) DAF thickeners. Either primary or secondary sludge can be pumped to these units, but currently they are only used for thickening secondary sludge. Thickened sludge (float), and the heavy solids that settle and are removed

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[1] It is estimated that 0.07 to 0.1 T/MGIF is grit, that is not being removed in the existing aerated grit chambers.

from the bottom of the tanks, can be pumped either to the anaerobic digestion system or directly to raw sludge dewatering. Subnatant from these units is returned to the high rate, activated sludge process. Operational data for the months of January through June 1980 are provided in Appendix A, Table A-3. Data analysis is provided in Appendix D.

Current practice is to pump waste secondary sludge 24 hours per day from the secondary system clarifiers at the maximum concentration possible (7,000 to 9,000 mg/l). Polymer is utilized for the following reasons:

1. Polymer usage is required to float and thicken effectively secondary waste-activated sludges containing iron salts.

2. Polymer usage maximizes solids recovery, thus minimizing the recirculation of solids via the subnatant stream that is returned to the high-rate, activated sludge process.

3. Polymer usage maximizes the solids loading rate that can be applied per unit of area per unit of tank, thus minimizing the number of DAF units required to be operated.

4. Polymer usage allows the subnatant stream to be used as a source of pressurized flow for the DAF thickener. If the subnatant stream could not be used, plant effluent would have to be utilized requiring extensive piping and increased operating cost.

Data analysis indicates that under this mode of operation:

1. The average thickened solids concentration that can be expected is 4.1 ± 0.3 percent.

2. The average thickened sludge volume per million gallons of plant influent flow is 1,644 ± 239 gallons. If secondary solids per million gallons of influent flow increase by approximately 15 percent (as discussed in Section 4), it is assumed that the thickened sludge volume will also increase proportionately by 15 percent to 1900 ± 239 gallons.

3. Dry polymer usage currently at 7 to 10 pounds per ton of dry secondary sludge solids is not expected to change in the future.

4. At current average plant influent flows, it is estimated that 7 DAF units are required to be operating. When secondary solids per million gallons of influent flow increase in the future it is estimated that 8 DAF units will be required.

5.2.1. Removal of Settled Solids

Not all secondary solids are capable of being treated by DAF. Those solids that are too heavy to float, settle to the bottom of the DAF unit and are conveyed by a chain and scraper mechanism to a sump located at the discharge end of each tank. Such solids were planned to be removed from each sump by use of a telescopic sludge valve; however, there are times when insufficient static head is available to make the heavy solids flow upward and out through the top of the telescopic valve. When this occurs, heavy sludge continues to accumulate in the bottom of the tank until it affects the performance of the unit, which then has to be taken out of service.

Although it is beyond the scope of this report to provide detailed solutions to operating problems, there are two possible methods of dealing with the problem that should be investigated:

1. Modifying each telescopic sludge valve to an air lift pump. Injecting about 3 to 5 cfm of air near the bottom of each valve would permit the movement of sludge through the telescopic valve.

2. Provide positive sludge removal through the use of pumps discharging into a common header located in the flotation thickener basement gallery.

5.2.2. Pressure Regulating Valve

Until recently, pressurized flow for DAF thickeners has been controlled through an air actuator and pneumatic pressure device. This automatic control system works very well when a high quality air supply is available and regular operator attention (two to three performance checks per 8-hour operating shift) is provided. When this actuator malfunctions, the float quality rapidly deteriorates, polymer usage increases, and consequently more DAF units have to be brought on-line to continue processing sludge.

This valve has been a continuous source of operating problems. Recently, the manufacturer of the existing flotation equipment has offered a new type of spring-operated pressure relief valve in place of the air actuated system. Figure 5-1 shows a cross-sectional view of the valve assembly.

This spring-operated pressure relief valve relies on an increase in fluid pressure to push the valve diaphragm against the spring. Since the spring has a set constant of compression, it permits the diaphragm to open just far enough to maintain the set pressure and allows flow to pass. This valve has been tested at one facility for over three years and has performed very well. It is recommended that the WWTF personnel consider use of this new type valve.

5.2.3. Spare Part Requirements

As a result of the overall financial constraints, spare parts inventory is almost non-existent. In order to keep the required number of DAF units in operation, several units have been cannibalized" for parts. Currently, this is not a critical problem because excess capacity for normal operation is available, but if allowed to continue this could lead to significant sludge processing problems.

5.3. ANAEROBIC DIGESTION SYSTEM OPERATION

The anaerobic digestion system is made up of several major subsystems: digestion tanks, sludge heating, digestion tank sludge mixing, gas collection and storage, and sludge piping and pumping capabilities. At the present time only a fraction of the raw sludge is pumped continuously to the digestion system. Operational data from January 1979 through June 1980 is provided in Appendix A, Table A-4.

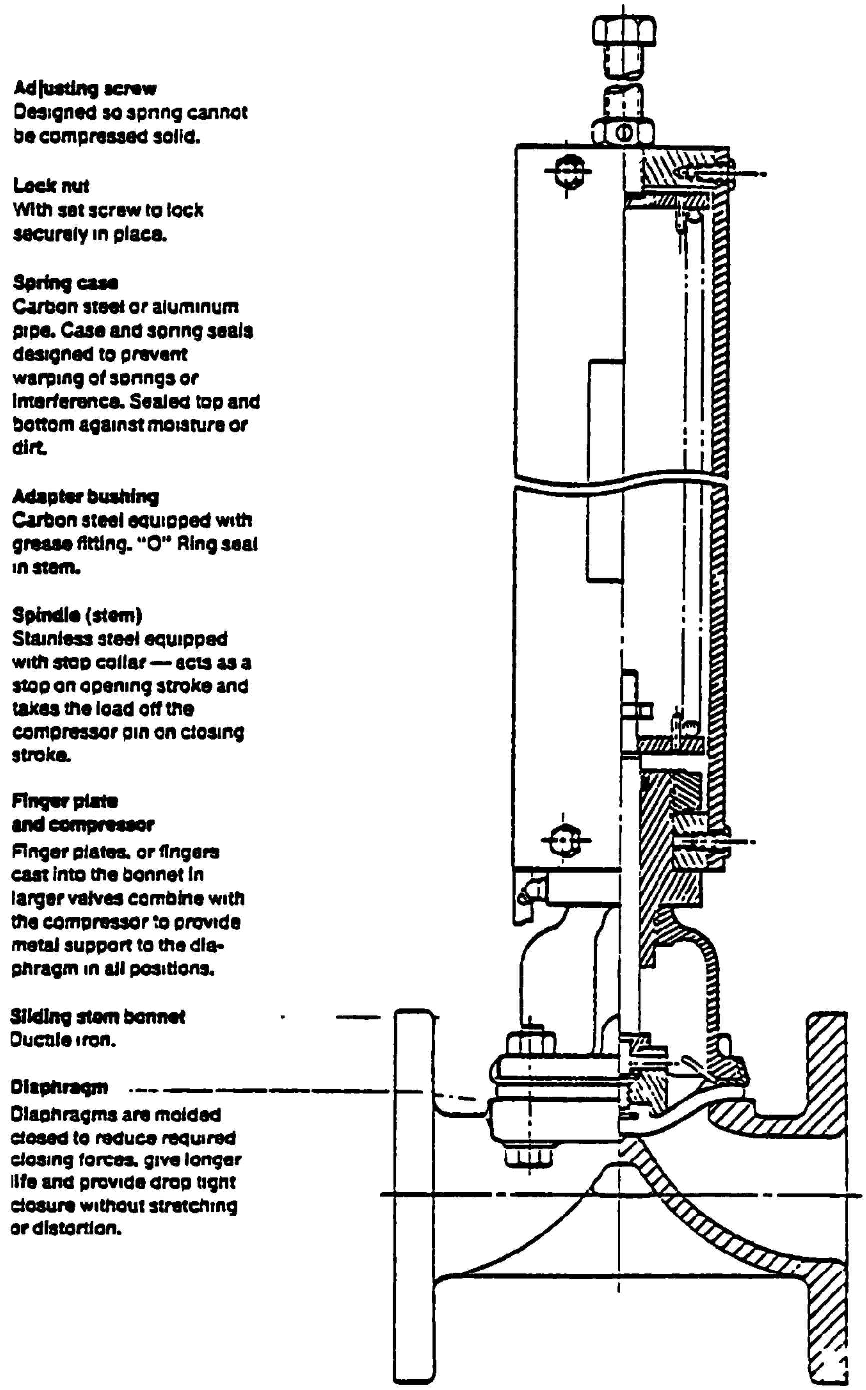

FIGURE 5-1. Cross-sectional view of spring-operated pressure relief valve

5.3.1. Digestion Tanks

There are 12 earth covered, concrete, digestion tanks. Each tank has an inside diameter of 84 feet and a theoretical liquid volume of 146,500 cubic feet (1,096,000 gallons). Though no longer usable, internal heating coils are still located inside each unit. Lithium chloride tracer studies (Appendix E) indicate that approximately 65 percent of the total actual volume--1.14 million cubic feet (8.54 million gallons)--is available. The remaining 35 percent is unusable due to such factors as (1) volume occupied by scum and grit and (2) inadequate mixing.

Scum--The scum thickness is measured on all tanks on a routine basis. The findings indicate that the thickness of the layer depends upon the operation of the digestion tank gas-mixing system (Section 5.3.3). When the gas-mixing system is not operational, the scum layer builds up rapidly to a thickness of several feet. When the system is in operation, very little scum accumulates.

Grit--On a routine basis, digestion system operations personnel probe the tank bottoms for grit buildup. Grit in the digesters has always been a problem. Until several years ago, digesters were systematically taken out of service for cleaning, but this is no longer done. Based on conversations with long-time plant operators, grit accumulates in each tank until it reaches a point of equilibrium. Best estimates are that 15 to 20 percent of each tank is occupied by grit. Using these assumptions, calculations in Appendix F indicate that for each million gallons of influent flow, 1.4 to 2.0 cubic feet of grit passes through the aerated grit chambers and is removed with the primary sludge.

5.3.2. Sludge Heating

There are six, double inlet, double outlet, 6 inch by 8 inch, tube and tube heat exchangers. The source of hot water is from two steam boilers and one hot water boiler, each fueled by digester gas with diesel standby.

Originally designed to transfer 3,000,000 BTU per hour per heat exchanger (18,000,000 BTU per hour total), current operation is only capable of 13,000,000 BTU per hour total. Refurbishing of the existing hot water boiler in 1981 will add another 3,000,000 BTU per hour to the total. In addition, the current on-going construction contract will tie in the existing steam boilers with new steam boilers elsewhere on the plant site. These new steam boilers will have an approximate total capacity of 17,000,000 BTU per hour, but how much could be used for sludge heating is unknown. Appendix G gives a detailed analysis of sludge heating capacity. Appendix G also analyzes system heat requirements for both summer and winter operation. Table 5-1 summarizes the results.

TABLE 5-1. SUMMARY OF ANAEROBIC SYSTEM HEAT REQUIREMENTS[a]

	Winter operation	Summer operation
For raw sludge addition in BTU's per MGIF[b]		
Primary sludge only	511,808	255,904
Secondary sludge only	636,353	272,723
Total primary and secondary sludge	1,148,161	528,627
For system radiation heat losses in BTU's per hour		
Roof	1,453,112	535,358
Walls	613,276	285,826
Floor	546,866	546,866
Sludge piping	582,400	467,200
Total radiation losses	3,195,654	1,835,250

[a] Calculations for data are presented in Appendix G.

[b] MGIF - million gallons influent flow.

Based on the information presented in Appendix G, there is currently 9,800,000 BTU per hour available for raw sludge addition during the winter months and 11,100,000 BTU per hour for raw sludge addition during the summer months. After refurbishing the existing hot water boiler, raw sludge heating capacity will increase to 12,800,000 BTU per hour and 14,100,000 BTU per hour, respectively.

5.3.3. Digestion Tank Sludge Mixing

Mixing in the anaerobic digestion tanks is primarily done using recirculated gas. Three 1,000 gallon per minute recirculation pumps are available for substitute or supplemental mixing.

All gas-mixing systems are of the type in which digester gas is removed from the tank, compressed, and discharged into the lower part of the digester through vertical 2-inch pipes (lances) supported from the tank roof. In some tanks, gas is discharged through only one pipe at a time, the point of discharge being changed to another discharge pipe either manually or automatically on a timed basis while others discharge through all points at once.

There are several conditions that prevent the existing gas mixing systems from adequately mixing each tank.

1. There is a significant amount of grit accumulating in each tank. Since grit has a specific gravity of 2.65, digestion gas-mixing systems cannot cope with these accumulations.

2. Since all tanks contain their old internal heating coils, which completely circle the inside of the tank, the walls act as barriers to the mixing currents thus allowing material to build up behind them and reduce useable tank volume.

3. Eight tanks (3, 4, and 7 through 12) were equipped with gas-mixing systems in 1960 that were intended to serve as scum breakers until they could be replaced by a permanent gas recirculation system.

Since 1978, several of these tanks have had newer systems installed, but they have been unable to operate continuously because of the lack of spare parts.

4. As a result of the overall financial constraints, spare parts for the gas-mixing systems have been difficult to obtain and one to two gas-mixing systems have been inoperative.

5.3.4. Gas Collection and Storage

Anaerobic digestion produces a gas with an energy value of roughly 600 BTU per cubic foot and the District has the potential of producing two million cubic feet per day (see Section 5.3.7 for analysis of gas production capabilities).

Visual inspection of the existing gas collection and storage system revealed that there is need for appropriate corrective action at least in the following instances.

1. Some parts of the gas piping system show extensive corrosion, to the extent that holes have developed in the piping, with duct tape wrapping used to prevent leakage.
2. Extensive corrosion on the bottom of the floating gas holder prevents the unit from being used to its fullest capacity. Approximately 20 percent of the useable volume has been lost.
3. Some gas safety equipment was inoperative.
4. Where operators must occasionally enter several locations where digester gas can build up. No gas alarm or portable gas meter has been provided for operator safety.

5.3.5. Sludge Piping and Pumping Capabilities

The anaerobic digestion system is provided with an extremely flexible sludge piping and pumping system. Appendix H contains a current sludge piping schematic that was developed as part of this study. A cursory review of existing line sizes and pump capabilities indicated that capacity is adequate to handle all sludge generated.

5.3.6. Volatile Matter Reduction

Due to the conversion of some of the biodegradable organic material (volatile solids) to methane (CH_4) and carbon dioxide (CO_2), anaerobic digestion of municipal wastewater sludge decreases the total solids. The amount of volatile material that can be reduced depends primarily on sludge type, digestion operating temperature, food to organism ratio, and length of time the sludge is kept within the digestion tank (sludge residence time). In addition, at similar digester operating temperatures and sludge residence times, a greater fraction of primary sludge will digest than waste activated sludge. Figure 5-2 shows a plot of percent volatile solids

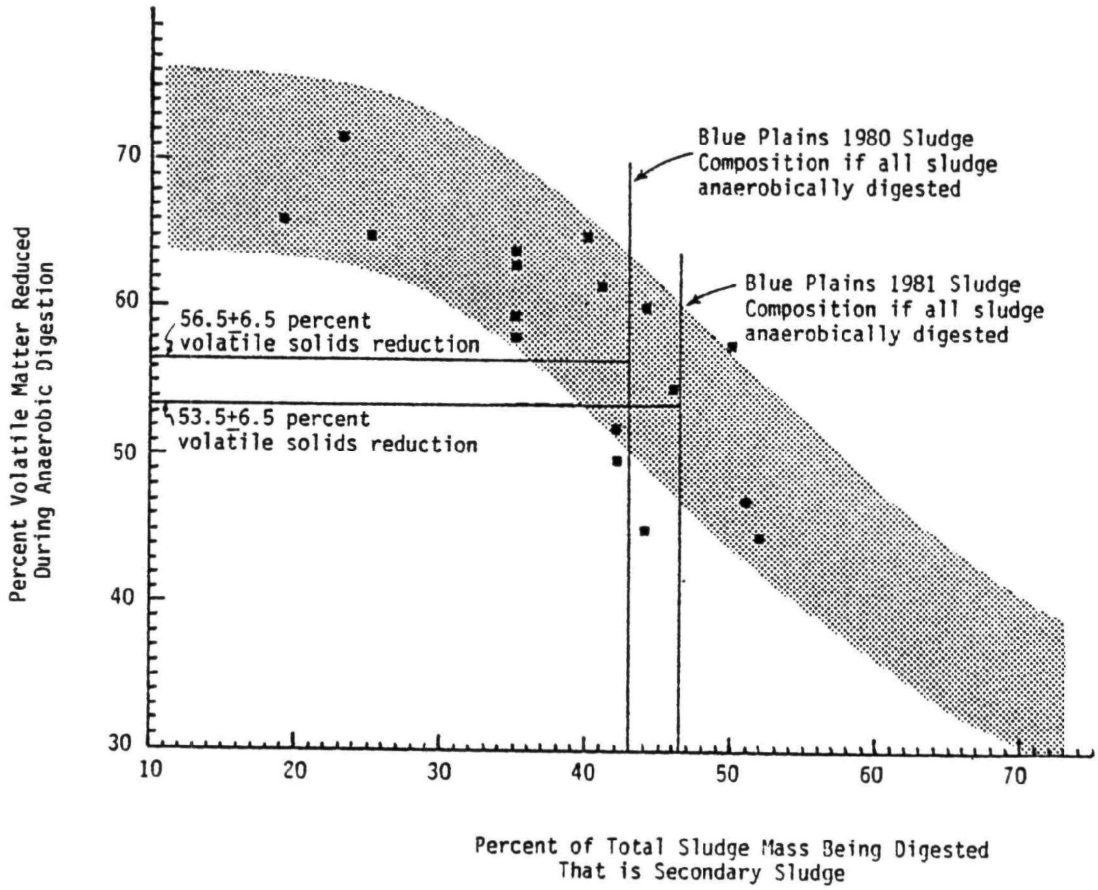

FIGURE 5-2. The effect of secondary sludge on volatile matter reduction at the blue plains anaerobic digestion facility[2]

[2]See Appendix I for development of data points.

reduction versus percent by weight of secondary sludge in the sludge mass being digested at digestion temperatures of 93 to 95°F and sludge residence times generally from 17 to 21 days. Each point on Figure 5-2 represents the average performance over a one-month period.

As the secondary sludge fraction of the total sludge mass increases (digestion temperature and sludge residence time held relatively constant), the percent volatile matter reduction decreases as would be expected (Figure 5-2). The curve drawn in Figure 5-2 is of "eye ball" fit constructed under the following assumptions.

- o The lack of accuracy involved in performing solids analysis plus the fact that each data point represented the average of one month of operating data dictated that a wide band curve incorporating a range would be required.
- o Realization that without data in the region of 0 to 20 percent on the horizontal axis, the curve would have to lie between 60 to 80 percent volatile matter reduction as the secondary sludge fraction of the total mass approached zero percent (USEPA, 1979).
- o Realization that without data in the region beyond 52 percent on the horizontal axis, the curve would have to lie between 30 to 40 percent volatile matter reduction as the secondary sludge fraction of the total mass approached 100 percent (USEPA, 1979).

Volatile matter reduction as a function of organic loading and hydraulic residence time are given in Figures 5-3 and 5-4, respectively. In evaluating the information shown in these figures, it must be remembered that not all sludge goes to anaerobic digestion, in fact not even the same ratio of primary to secondary sludge is maintained. The amount and ratio pumped to digestion is presently controlled by the raw and anaerobically digested sludge dewatering operations.

Organic loading over the range indicated in Figure 5-3 had little or no effect on volatile matter reduction. Analysis of the raw data indicates that at the higher organic loadings and higher volatile matter reductions, the ratio of primary sludge to secondary sludge was considerably higher than the normal 50:50 to 60:40 split. These results would be expected since Figure 5-2 showed that a greater fraction of primary sludge digested than secondary sludge.

Evaluation of the raw data in Figure 5-4 indicates that the four points beyond 22 days are for time periods in which the primary:secondary sludge ratio was considerably higher than the normal 50:50 to 60:40 split. If these data points were removed, the data would indicate no effect of hydraulic detention time on volatile matter reduction over the range indicated. This would imply that a primary:secondary sludge ratio of 50:50 to 60:40 could operate at less than 16 days hydraulic detention and still expect 45 to 60 percent volatile matter reduction.

Based on the curve developed in Figure 5-2 and the data presented in Figures 5-3 and 5-4, the following can be said about volatile matter reduction using the existing anaerobic digestion system. (Assuming hydraulic detention times of 16 to 17 days, operating temperatures of 93°F to 95°F, and organic loadings of 0.12 to 0.14 pounds volatile matter per cubic foot per day.)[3]

- Digestion at the current secondary sludge to total sludge mass generation ratio of 0.43 should provide a 56.5 ±6.5 percent volatile matter reduction. This would give an overall reduction in solids of 0.256 ± 0.042 T/MGIF.
- Digestion at the future secondary sludge to total sludge mass generation ratio of 0.465, should provide a 53.5 ± 6.5 percent volatile matter reduction. This would give an overall reduction in solids of 0.255 ± 0.042 T/MGIF.

5.3.7. Gas Production

Anaerobic digestion produces a gas with an energy value of approximately 600 BTU per cubic foot. As indicated in Appendix I, in order for the existing heat exchangers to operate at their maximum capacity, the boilers would require 810,000 cubic feet of digester gas per day. In addition, the District WWTF is negotiating with the Naval Research and Development Center to sell them excess digester gas at $0.95 per 1,000 cubic feet.

[3] Data analysis for volatile matter reduction is given in Appendix I.

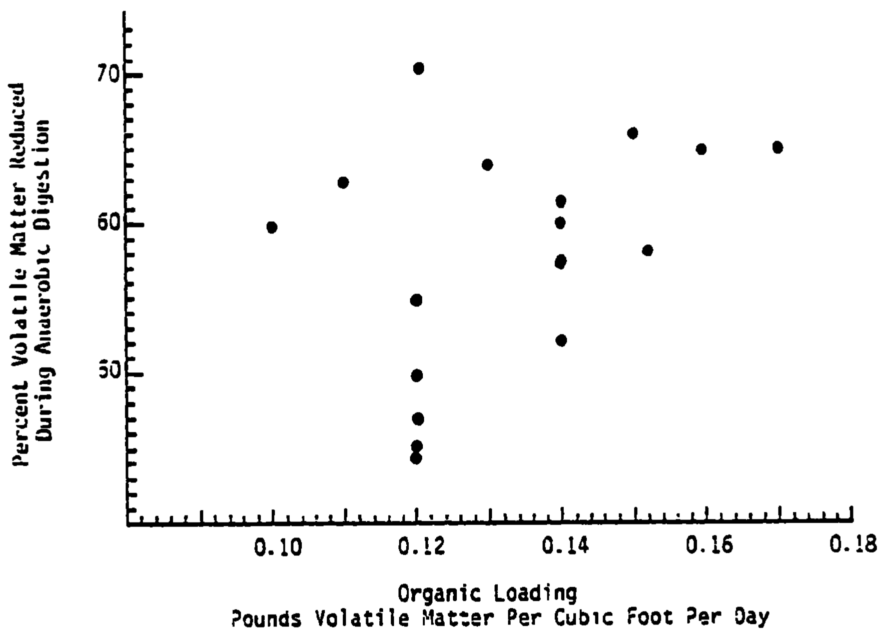

FIGURE 5-3. The effect of organic loading on volatile matter reduction at the District of Columbia anaerobic digestion WWTF[4]

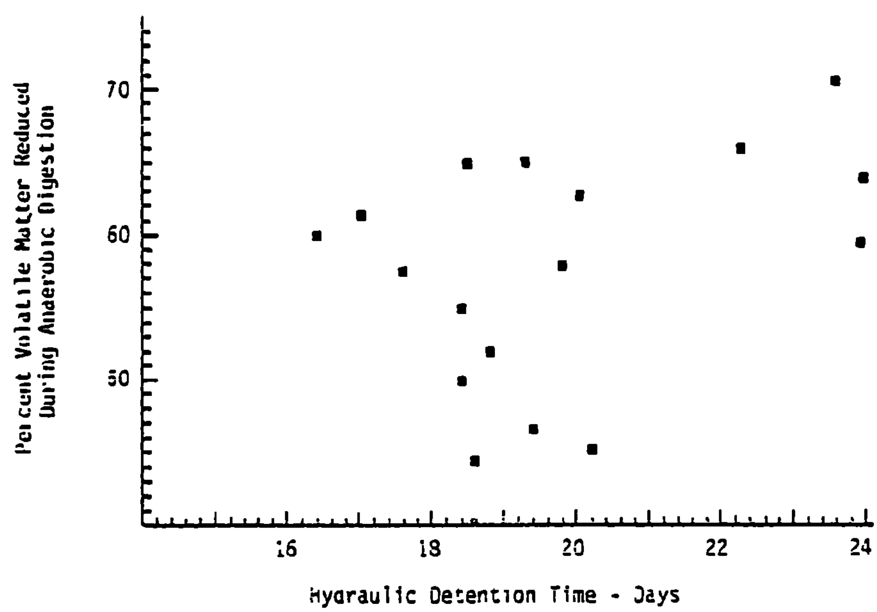

FIGURE 5-4. The effect of hydraulic detention time on volatile matter reduction at the District of Columbia anaerobic digestion WWTF[3]

[4] See Appendix I for development of data points.

Figures 5-5, 5-6, and 5-7 show operating data on gas production per pound of volatile matter reduced as a function of feed solids concentration, and volatile matter loading. All data indicate that significant amounts of gas production can be lost as a result of digestion system process stress, possibly caused by either poor mixing or ammonia toxicity.

Gas production per unit of volatile matter reduced declined with increasing digester feed solids concentration (Figure 5-5). This decline may be attributed to the existing systems inability to mix the solids at higher solids concentrations. Insufficient mixing allows concentration gradients to build up which tend to disrupt the biological process. Two possible solutions would be (1) increase the mixing intensity or (2) operate the system at a lower feed solids concentration.

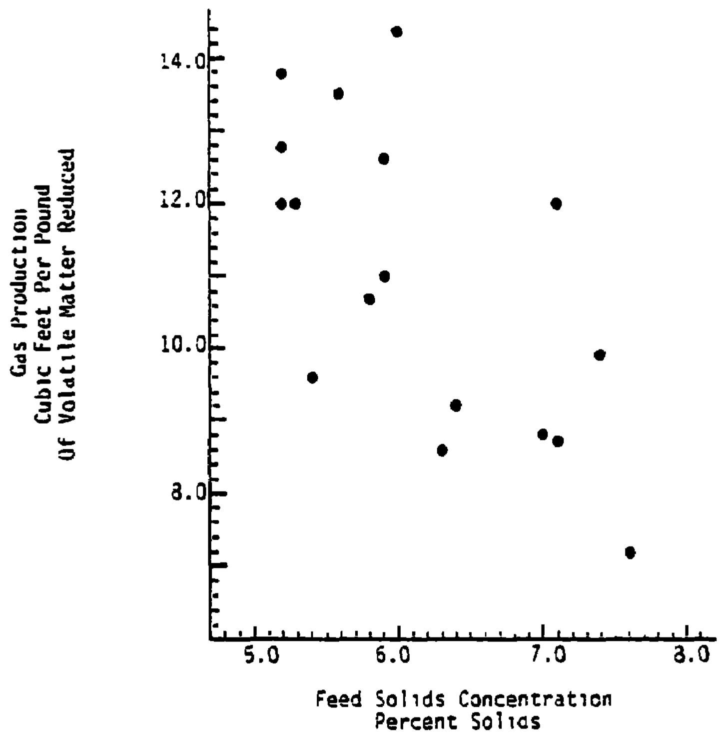

FIGURE 5-5. The effect of feed solids concentration on anaerobic digestion gas production at the District of Columbia digestion WWTF[5]

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[5] See Appendix I for development of data points.

The decline may also be attributed to ammonia toxicity. Ammonia nitrogen levels in the digestion tanks vary from 1,200 to 1,500 milligrams per liter (mg/l). It is known that ammonia nitrogen starts to affect the digestion process adversely above concentrations around 1,400 to 1,500 mg/l. Therefore, it may be concluded that reduced gas production at the higher feed solids concentration is a reflection of the higher potential for ammonia nitrogen toxicity.

The solution to ammonia toxicity is dilution. This dilution would be accomplished by proper blending of the thickened primary and secondary sludge. It is desirable to digest as concentrated a sludge as possible and still maximize gas production. From Figure 5-5, a value of six percent meets this criterion. To meet the six percent level, the primary sludge mass should not be more than 65 percent of the total sludge mass pumped to digestion.

Gas production per unit of volatile matter reduced declined with increased volatile matter loading (Figure 5-6). The current operational practice is to maintain a constant liquid volume flow to digestion, but to vary the primary-secondary sludge mixture.

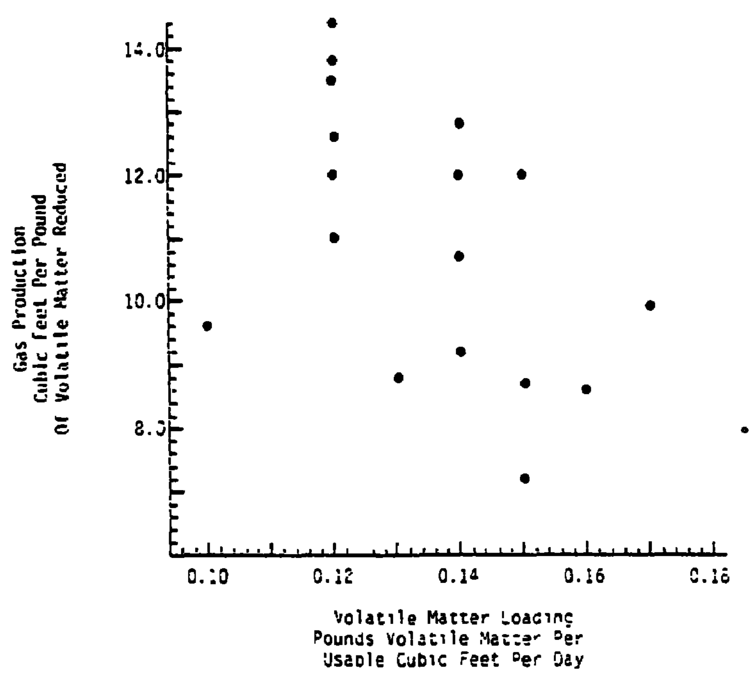

FIGURE 5-6. The effect of volatile matter loading on anaerobic digestion gas production at the District of Columbia digestion WWTF[6]

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[6] See Appendix I for development of data points.

Increasing the percentage of primary sludge means more volatile matter per unit of flow, thereby increasing the volatile matter loading on the digestion system. Therefore, the higher volatile matter loadings indicated in Figure 5-6 are due to a high percentage of primary sludge in the incoming flow. This decline may again be due to poor mixing at the higher feed solids concentration that would develop with the greater percentage of primary sludge being pumped to the system. It is desirable to maintain the highest volatile matter loading possible and still maximize gas production. From Figure 5-6, a value between 0.12 to 0.14 meets this criterion.

5.4. ELUTRIATION SYSTEM OPERATION

Elutriation (washing) of the anaerobically digested sludge is practiced to reduce chemical conditioning requirements (reduced operating cost) in the vacuum filter operation. River water is used as the source of washwater.

The elutriation system consists of two 35-foot wide by 70-foot long tanks, each equipped with sludge removal and surface skimming equipment. The tanks are operated in series. Anaerobically digested sludge is mixed with wash water before entering the first tank. The sludge that settles to the bottom is removed and pumped to the second tank where it is again mixed with wash water. Settled sludge from the second tank is pumped to the vacuum filter. Part of the elutriation tank overflow flows by gravity to the primary clarifiers and the remainder is pumped to the gravity thickeners. Polymer is added to improve solids concentration and capture.

Figures 5-7 through 5-9 show the effect of several variables on the percent solids concentration of the elutriation underflow pumped to the vacuum filters. The data presented in these figures are included in Appendix A, Table A-5 and in Appendix J. As will be demonstrated in Section 5.5, it is important to maintain this solids concentration as high as possible.

Figure 5-7 indicates that a feed solids concentration of at least 2.7 percent is required to develop a five percent underflow concentration. Because there is no digester supernatant, the solids concentration out of the digesters is controlled by the solids concentration into the digesters. Assuming an average 53 percent volatile matter reduction (Section 5.3.6), the feed solids concentration into the digester must be a minimum of 4.5 to 5 percent. This also implies that the primary:secondary sludge mixture has to be a minimum of 1:6.

The total hydraulic flow to the elutriation system is composed of digested sludge flow and washwater flow, which currently totals 2.2 to 2.4

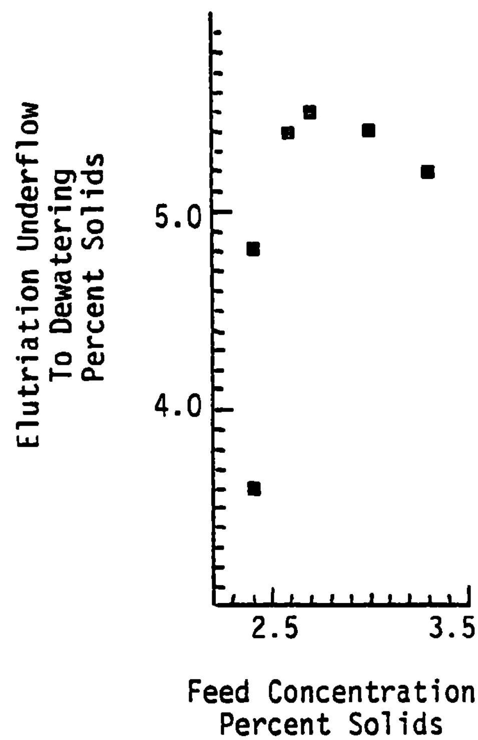

FIGURE 5-7. The effect of elutriation feed solids concentration on elutriation underflow solids concentration[7]

[7] See Appendix J for development of data points.

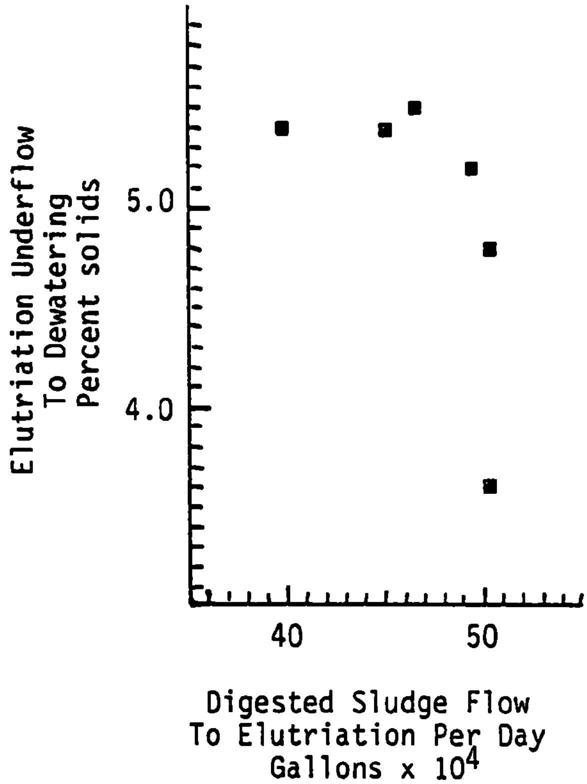

FIGURE 5-8. The effect of digested sludge flow rate on elutriation underflow solids concentration[7]

million gallons per day. Higher total flows are not possible because of the inability to remove elutriation overflow at higher flow rates. Preliminary analysis of elutriation underflow solids concentration versus total flow rate showed no relationship. Evaluation of the digested flow component (Figure 5-8) indicates that elutriation underflow solids concentration is quickly and significantly reduced at digested flow rates over 450,000 to 470,000 gallons per day.

[7] See Appendix J for development of data points.

The addition of washwater improves the compaction capabilities of sludge; however, according to Figure 5-9, washwater usage beyond 4.5 to 4.7 times the incoming digested sludge flow rate is of no value. Because of the costs to pump washwater from the river and treat it after use, it is best to minimize the quantity of washwater consumed for elutriation.

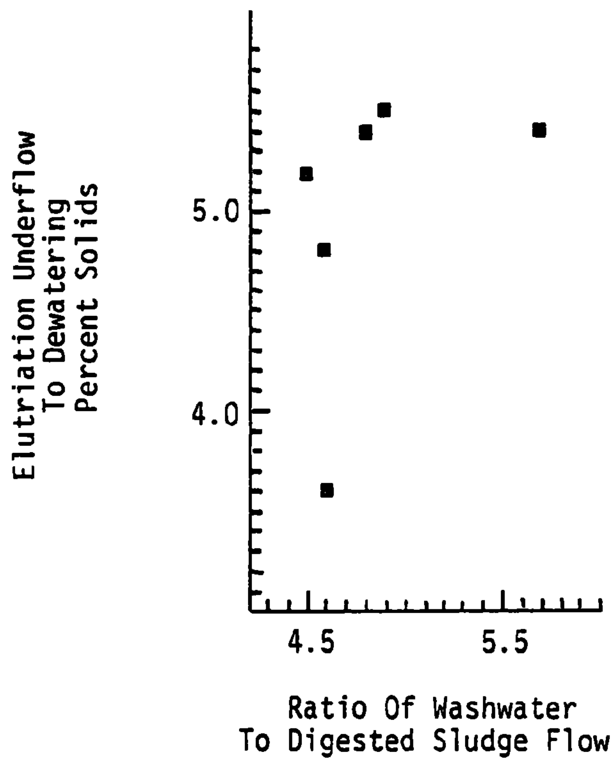

FIGURE 5-9. The effect of washwater on elutriation underflow solids concentration[8]

[8] See Appendix J for development of data points.

5.5. ANAEROBIC SLUDGE DEWATERING OPERATION

5.5.1. Solids Generation

Currently, anaerobically digested sludge is dewatered on drum-type rotary vacuum filters after conditioning by elutriation and the addition of ferric chloride. The addition of ferric chloride adds solids to the total sludge mass.

The amount of ferric chloride added is governed by several factors: the operators visual evaluation of cake release from the drum and cake solids content, both of which will improve with increasing ferric chloride addition; and management's attempt to keep ferric chloride addition at a minimum level and still obtain acceptable results. Although existing data are inconclusive, there are indications that some relationship is exerted on the ferric chloride addition by the filter feed solids concentration (Figure 5-10) and by the percent secondary sludge in the total sludge mass (Figure 5-11).

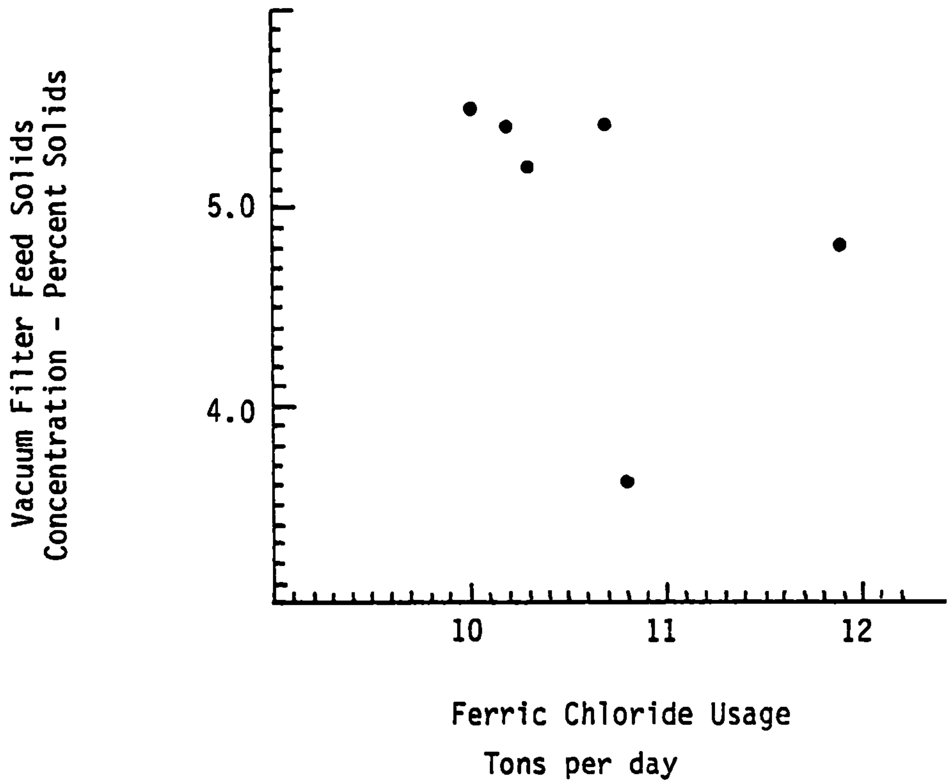

FIGURE 5-10. The effect of ferric chloride addition on filter feed solids concentration[9]

[9] Filter operates 24 hours per day. See Appendix K for development of points.

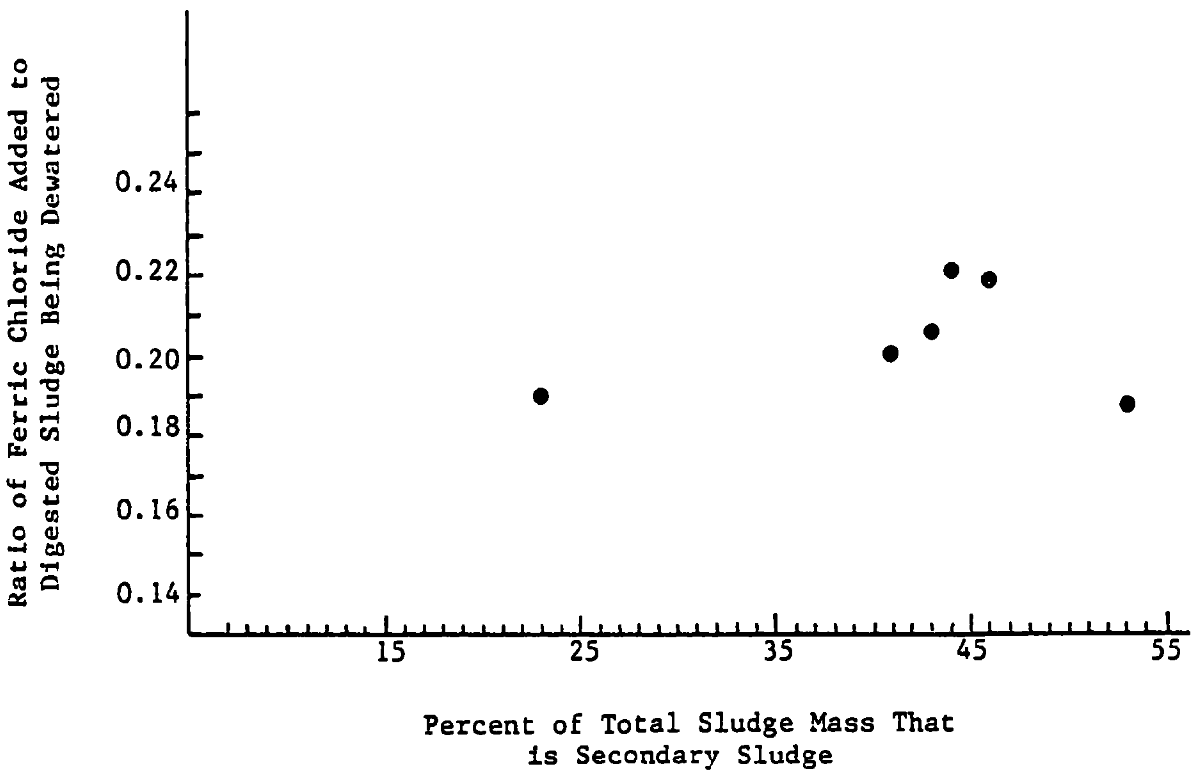

FIGURE 5-11. The effect of secondary sludge mass on ferric chloride addition for vacuum filter dewatering of anaerobically-digested sludge[10]

Since a positive relationship for either Figure 5-10 or 5-11 cannot be developed at this time, the amount of solids contributed by ferric chloride addition was calculated on a more empirical basis, as described in Appendix K. Based on the analysis conducted, the following can be stated about increased solids production resulting from ferric chloride addition for dewatering digested sludge.

[10] See Appendix K for development of data points.

1. At current conditions for the sludge being anaerobically digested, each million gallons of influent flow generates an additional 0.043 ± 0.009 tons of solids.

2. At future conditions for the sludge being anaerobically digested, each million gallons of influent flow would generate an additional 0.049 ± 0.009 tons of solids.

5.5.2. Cake Dryness

Final disposal of digested sludge requires truck transport to a final disposal site. Minimizing the volume of digested sludge produced, minimizes the transporation cost.

Figure 5-12 indicates that within the range evaluated, increasing feed solids concentration produced an increasingly drier filter cake. (Note: Feed solids concentration is the same as that in the elutriation underflow.) Therefore, from the standpoint of the dewatering operation, the digested sludge should be maintained at the highest possible feed solids concentration.

5.6. RAW SLUDGE DEWATERING OPERATIONS

5.6.1. Solids Generation

Currently, raw sludge is dewatered on cloth belt type, rotary vacuum filters. Sludge conditioning consists of addition of both ferric chloride and lime, which adds substantially to the total sludge mass.

No analysis was conducted to evaluate if any relationship existed for predicting ferric chloride and lime usage as a function of raw sludge characteristics, (for example, feed solids content, or secondary fraction of the total sludge mass). Lime is currently added in excess of conditioning requirements to maintain a high pH to meet sludge trenching requirements. In the near future when trenching is discontinued, lime requirements would be expected to decrease by at least 50 percent, which would affect the amount of ferric chloride utilized.

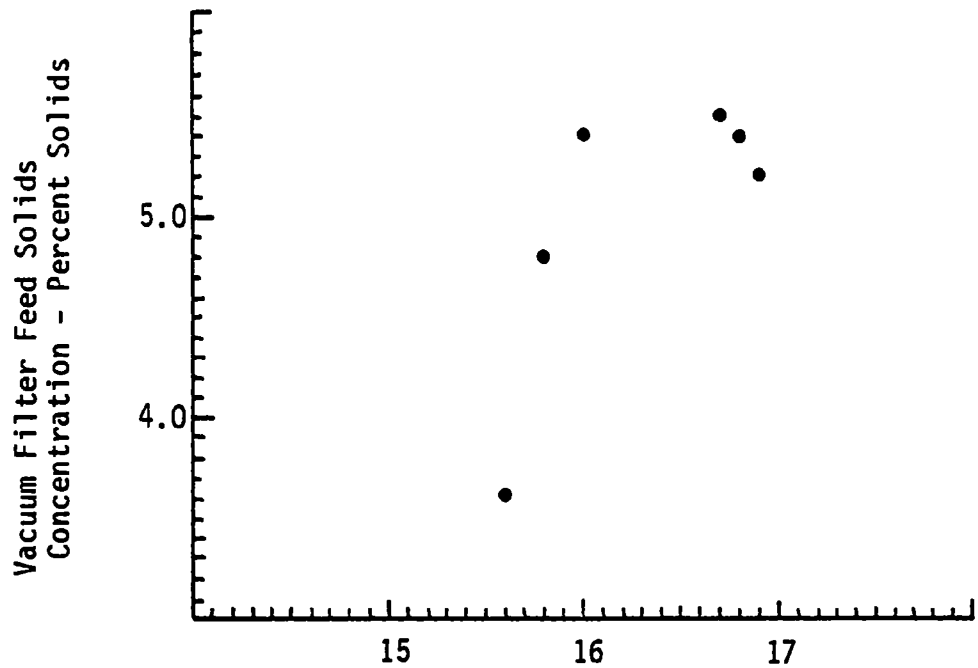

FIGURE 5-12. The effect of feed solids concentration on vacuum filtration of anaerobically digested sludge[11]

The amount of solids contributed by ferric chloride and lime in dewatering raw sludge is calculated in Appendix L. The results of these calculations are as follows:

1. At current conditions, each million gallons of influent flow generates an additional 0.049 ± 0.015 tons of solids from ferric chloride addition and 0.223 ± 0.042 tons from lime addition.

2. In the future, even if raw sludge was not to be trenched but raw sludge still needed to be dewatered, each million gallons of influent flow would generate an additional 0.044 ± 0.015 tons of solids from ferric chloride addition and 0.119 ± 0.040 tons from lime addition.

3. In the future, even if raw sludge was not to be trenched but raw primary sludge still needed to be dewatered, each million gallons of influent flow would generate an additional 0.024 ± 0.009 ton of solids from ferric chloride addition and 0.064 ± 0.022 tons from lime addition.

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[11] See Appendix K for development of data points.

4. In the future, even if raw sludge was not to be trenched but raw secondary sludge still needed to be dewatered, each million gallons of influent flow would generate an additional 0.021 ± 0.007 tons of solids from ferric chloride addition and 0.055 ± 0.018 tons from lime addition.

5.6.2. Cake Dryness

As with digested sludge, final disposal of raw sludge requires truck transport to a final disposal site. Minimizing the volume of raw sludge is again important to minimizing the transportation cost.

Figure 5-13 indicates that within the range evaluated, increasing feed solids concentration produced an increasingly drier cake. Therefore from the standpoint of the dewatering operation, the raw sludge should be maintained at the highest possible feed solids concentration. In the future when lime and ferric chloride additions are reduced, it is believed that the highest cake solids will still be produced from the highest feed solids concentration, but that the cake dryness will not be as high as currently developed for a given feed solids concentration.

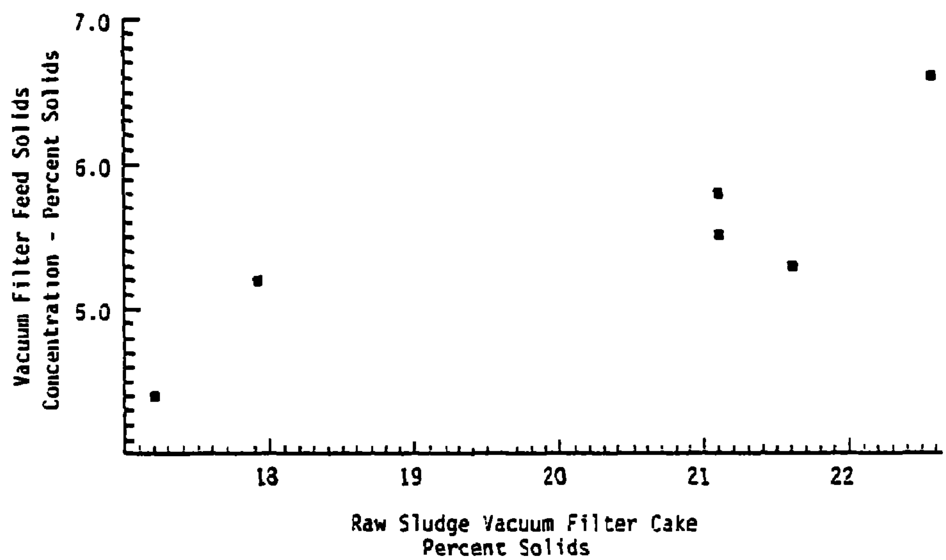

FIGURE 5-13. The effect of feed solids concentration on vacuum filtration of raw sludge [12]

[12] See Appendix L for development of data points.

Figure 5-14 indicates that decreasing amounts of secondary solids in the mixture produce drier dewatered cake. This would be expected since primary solids are gritty and fibrous, which compact better than gelatinous secondary solids.

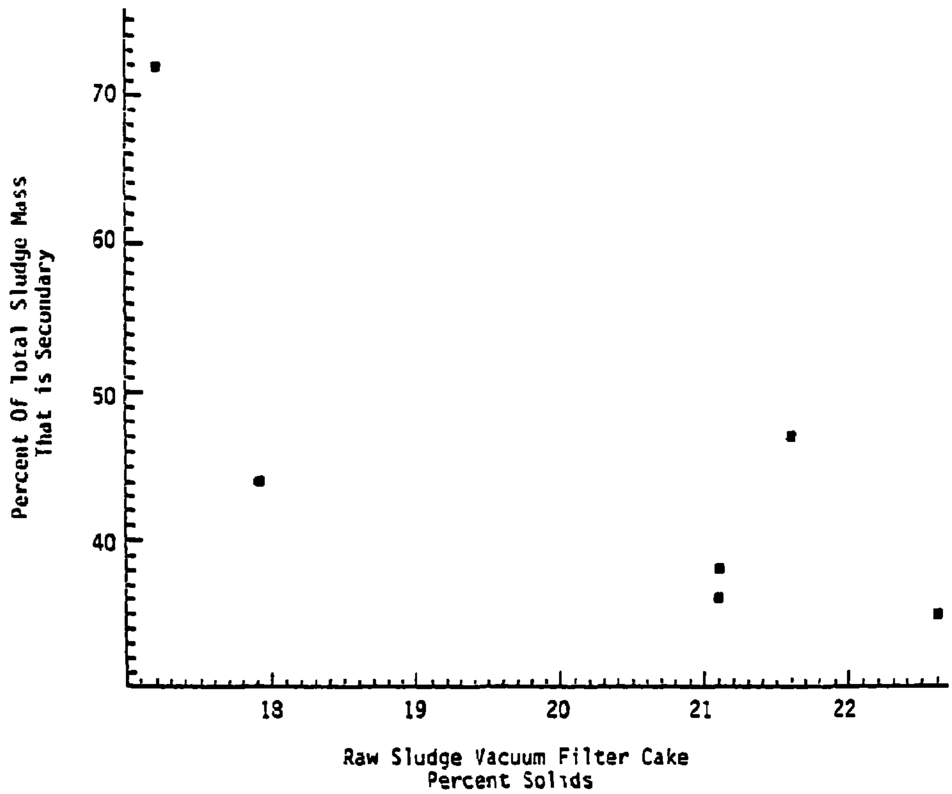

FIGURE 5-14. The effect of secondary sludge on vacuum filtration of raw sludge[13]

5.7. FINAL SLUDGE DISPOSAL

Currently, several methods of final disposal for the sludge generated at the District of Columbia's WWTF are used:

o Dewatered anaerobically digested sludge is trucked and disposed of on land for approximtely $15 per wet ton.

o Approximately 100 wet tons per day of dewatered raw sludge is trucked to Beltsville at an approximate trucking cost of $30 per wet ton.

o Between 100 to 150 wet tons per day of dewatered raw sludge is trucked to MIT for composting. The approximate trucking cost is $30 per wet ton.

[13] See Appendix L for development of data points.

- The balance of the dewatered raw sludge is trenched. Hauling and disposal cost are approximately $35 per wet ton.

5.8. SUMMARY

The discussions of changes in solids production per MGIF caused by the various processing operations are summarized in Table 5-2. Table 5-3 presents the various limitations noted for the processing steps.

TABLE 5-2. SUMMARY OF ADDITION (REDUCTION) IN SOLIDS PRODUCTION DUE TO VARIOUS SOLIDS PROCESSING STEPS

Processing operation	Current sludge generation, dry tons (pounds) per million gallons of influent flow	Future sludge generation dry tons (pounds) per million gallons of influent flow
Gravity Thickening	None	None
Dissolved air flotation thickening	0.001 (2)	0.002 (3)
Anaerobic digestion	0.256 ± 0.042 ((512 ± 84))	0.255 ± 0.042 ((510 ± 84))
Elutriation	None	None
Digested sludge dewatering		
o Ferric chloride	0.043 ± 0.009 (86 ± 18)	0.049 ± 0.009 (98 ± 18)
Raw sludge dewatering		
o Ferric chloride	0.049 ± 0.015 (98 ± 30)	0.044 ± 0.015 (88 ± 30)
o Lime	0.223 ± 0.042 (446 ± 84)	0.119 ± 0.040 (238 ± 80)

TABLE 5-3. SUMMARY OF DISTRICT OF COLUMBIA'S WWTF SLUDGE MANAGEMENT PROCESS LIMITATIONS UNDER CURRENT OPERATION

Process	Limitations	
Gravity thickening	Average thickened primary sludge concentration is 7 0 ± 0 5 percent	
	Average thickened primary sludge volume per MGIF is 1,236 ± 218 gallons	
	Average primary solids production per MGIF is 0.377 ± 0 060 tons (754 ± 120 pounds) The average volatile content is 71 ± 5 1 percent	
Dissolved air flotation thickening	Average thickened secondary sludge concentration is 4 1 ± 0 3 percent	
	Average current thickened secondary sludge volume per MGIF is 1,644 ± 239 gallons; in 1981, it will increase to 1,900 ± 239 gallons	
	Average secondary solids production per MGIF is 0 284 ± 0.034 tons (568 ± 68 pounds) with an average volatile content of 65 percent; in 1981, it will increase to 0 328 ± 0 034 tons (656 ± 68 pounds) with an average volatile content of 63 9 ± 2.1 percent	
Anaerobic digestion		
o Tankage	Approximately 65 percent of the total volume available or 1,142,458 usable cubic feet (8,545,587 usable gallons)	
o Heating capacity	After meeting system radiation, losses at a 95°F operating temperature available heating capacity is currently 9,800,000 BTU per hour during the winter and 11,100,000 BTU per hour during the summer In 1981, this will increase to 12,800,000 and 14,100,000 BTU per hour, respectively	
o For maximum gas production	Maximum feed solids concentration to digestion not to exceed six percent solids.	
	Maximum volatile matter loading not to exceed 0 14 pounds per usable cubic feet per day	
Elutriation	Digested sludge flow rate to system not to exceed 490,000 gallons per day.	
	Feed solids concentration to be a minimum of 2 7 percent	
Digested sludge dewatering	Maximum percent cake solids achievable on a consistent basis 16 to 17 percent	
Raw sludge dewatering	Maximum current percent cake solids achievable on a consistent basis 21 to 22 percent In the future, this may decrease to 19 to 20 percent	

SECTION 6

IMPROVING SLUDGE MANAGEMENT OPERATION USING ANAEROBIC DIGESTION

Table 6-1 summarizes expected sludge quantities at the current average plant influent flow rate of 334 million gallons per day. The purpose of this study was to evaluate the existing sludge processing operation with the intent of determining if a new two-stage, mesophilic-thermophilic anaerobic digestion process could be applied to digest all sludge production using the existing equipment in conjunction with a minimal capital expenditure. If possible, then sludge processing and disposal cost at Blue Plains would be reduced by more than 50 percent. In addition, the scope of work included recommendations to improve the current anaerobic digestion operation and a brief evaluation of the use of thermophilic digestion. This section of the report will discuss these three objectives in the following order:

- o Existing mesophilic system operation
- o Evaluation of thermophilic system option
- o Evaluation of mesophilic-thermophilic system option.

TABLE 6-1 EXPECTED PRIMARY AND SECONDARY SLUDGE QUANTITIES AT CURRENT 334 MILLION GALLON PER DAY INFLUENT FLOW[a]

	Gallons of sludge per day			Pounds of sludge per day			Volatile pounds sludge per day		
	Minimum	Average	Maximum	Minimum	Average	Maximum	Minimum	Average	Maximum
Primary sludge	340,012	412,824	485,636	211,756	251,836	291,916	156,347	178,802	207,260
Secondary sludge	554,774	634,600	714,426	196,392	219,104	241,816	125,691	140,22[illegible]	154,7[illegible]
TOTALS	894,786	1,047,424	1,200,062	408,148	470,940	533,732	276,038	319,03[illegible]	362,022

[a] Data calculated from Table 5-3

6.1. EXISTING MESOPHILIC SYSTEM OPERATION[1]

Evaluation of an existing anaerobic digestion process system for the purpose of improving operations begins at the "end and works towards the front." This is done to ensure that any individual process constraints, which normally affect the output of the preceding process, can be identified and incorporated into the operation of the preceding process. In addition, the evaluation will be made from the viewpoint that all the sludge is to be processed through the present system. It is believed that this is the only impartial way to compare all three digestion processes.

6.1.1. Ultimate Disposal

Ultimate disposal of anaerobically digested sludge is on the land. It is desirable to have the driest cake possible to minimize the transportation costs.

6.1.2. Digested Sludge Vacuum Filter Operation

The constraint from ultimate disposal is to produce the driest sludge cake possible. From the viewpoint of dewatering management, it is important to minimize chemical additions (chemical cost and extra solids). In Section 5.5, the data presented indicated that maximum cake solids and minimum chemical usage occurred at the highest feed solids concentration. The data in Section 5.5 also indicated that with proper operation of the elutriation system, a minimum of five percent feed solids concentration could be obtained on a regular basis. This would result in an average 16.5 percent cake solids concentration.

Current Operation--

There are four old drum vacuum filters, each with 500 square feet of filtering area. In 1979 when the secondary to total sludge mass ratio of the sludge being digested was between 0.4 to 0.5, the average calculated yield was 2.41 pounds per square foot per hour. Assuming that all four

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[1] All supporting calculations for Section 6.1 are given in Appendix M.

units were operating at the average yield, 115,680 pounds of anaerobically digested sludge could be dewatered. At an average 16.5 percent cake solids there would be 390.5 wet tons of sludge per day needing disposal. At the present, the limited capacity of these facilities is a major constraint in processing sludge through the digestion system. This will change shortly with the start-up of six new and larger filters.

Future Operation--

Sometime in early 1981, six new cloth belt vacuum filters will become operable for anaerobic sludge dewatering and the four existing units will be abandoned. Each new unit will have 600 square feet of filtering area and it is assumed the same yield. Under these conditions the digested sludge dewatering capacity would increase to 208,224 pounds per day. With these new filters in operation, dewatering capacity will no longer be the rate-limiting process. In fact, elutriation will be the rate limiting process and will limit digested solids requiring dewatering to 142,000 pounds per day (479 wet tons to disposal), which would mean only four of the six filters would need to be operated at any one time.

It should be noted that if all primary and secondary sludge could be mesophilically digested, 10 of the 15 vacuum filters currently being used to dewater raw sludge could be used to dewater digested sludge, as piping and valving exits for such a configuration.

6.1.3. <u>Elutriation</u>

The constraints from the dewatering operation are: (a) to produce a solids concentration (elutriation underflow solids concentration) of at least five percent solids; and, (b) the maximum amount of sludge capable of being dewatered in early 1981 is 208,224 dry pounds per day.

In Section 5.4, the data presented indicated that in order to achieve a constant five percent solids concentration in the elutriation underflow:

- o The flow rate to the elutriation system from the digestion system should not exceed 490,000 gallons per day.
- o The solids concentration of the influent digested sludge stream to elutriation had to be a minimum of 2.7 percent.

With the start-up of the new digested sludge dewatering operation, the existing elutriation system would become the next rate-limiting process, limiting sludge processing to 490,000 gallons per day (40 to 55 percent of the total sludge flow).

Consideration should be given to using some of the spare dissolved air flotation tanks as elutriation tanks. Eight of the spare tanks would provide an additional 880,000 gallons per day elutriation capacity under the constraints specified. This additional capacity would allow the elutriation system to process all sludge generated. Appropriate charges would enable pumping digested sludge and washwater to the flotation tanks.

6.1.4. Anaerobic Digestion

The constraints from the present elutriation facilities are:

(a) Digested solids concentration from the digesters to be a minimum of 2.7 percent.

(b) The flow rate from the digestion system cannot exceed 490,000 gallons per day.

(c) The maximum sludge mass not to exceed 208,224 dry pounds per day.

From the viewpoint of digestion management it is important to maximize flow rate, solids reduction, and gas production within the constraints stated. In Section 5.3, the data presented indicated the following additional operating constraints:

(d) Approximately 35 percent of the existing tank volume--1,757,856 cubic feet (13,148,763 gallons)--is unusable, usable tank volume is 1,142,606 cubic feet (8,546,693 gallons).

(e) Available sludge heating capacity after meeting system radiation heat loss requirements is 9,800,000 BTU per hour during the winter and 11,100,000 BTU per hour during the summer. In 1981, this will increase to 12,800,000 and 14,100,000 BTU per hour.

(f) Maximum gas production achievable on a consistent basis required that:

- o Maximum feed solids concentration to digestion not to exceed six percent solids
- o Maximum volatile matter loading not to exceed 0.14 pounds per usable cubic foot per day

(g) Existing data indicate that hydraulic residence times of 16 days present no problem over a wide range of feed combinations. Lower detention times may be possible, but no operating data are available.

Since all digestion tanks are being mixed and no supernatant is removed, the flow rate out of the digestion tanks should be approximately equal to the flow into the tanks. The flow rate through the tanks is governed by one of three constraints: the elutriation system, sludge heating, and vacuum filtration capacity.

The calculations on sludge heating capabilities indicate that the existing heat exchangers are capable of heating 683,700 gallons per day to 95°F during the 1980 winter and 1,549,000 gallons per day to 95°F during the summer of 1981. When the new hot water boiler is put in service during the summer of 1981, the winter capacity will increase to 893,000 gallons per day and 1,967,000 gallons per day during the summer.

It should be noted that in order to process all sludge through digestion, an additional 4,400,000 BTU per hour would be required during the winter months. As explained in Appendix G, 17,000,000 BTU per hour of new low pressure (9 psi) steam capacity is being interconnected with the existing sludge heat exchanger steam boilers. If the capacity is available, it may be possible to place steam lines up to the top of the digestion tanks and inject the steam directly into the digesters through the roof.

Based on a minimum 16-day hydraulic detention time and 65 percent usable digester volume, the flow rate would be 534,000 gallons per day. If the grit problem was solved and all existing tank capacity became available, then the flow rate would be 821,700 gallons per day.

It may be possible to operate this mesophilic system at a lower hydraulic detention time, but operating data do not currently exist to verify this.

Both heating capacity and hydraulic detention time flow rate maximums are above the elutriation operation 490,000 gallon per day constraint. Therefore, total raw sludge flow rate through the existing anaerobic digestion tanks should be and is limited to 490,000 gallons per day under the present circumstances.

In addition to hydraulic flow rate restrictions on the digestion tanks, volatile matter loading requirements must also be satisfied. Available data indicate that the system can be successfully operated at volatile matter loading ratios of 0.16 to 0.17 pounds volatile matter per usable cubic foot per day, though gas production seems to deteriorate over 0.14. Based on 0.16 loading and 65 percent usable digester volume, the total amount of volatile solids pumped to the system would be 183,000 pounds per day. If the grit problem was solved and all existing tank capacity became available, then the amount would be 281,000 pounds per day.

If all the existing digesters were utilizing full capacity, then at least four more tanks identical to the existing 12 would be required to meet volatile matter loading requirements.

6.1.5. Primary - Secondary Flow Rates To Digestion

The ratio of thickened primary to thickened secondary sludge volumes is 0.65:1. At the current time, the ratio being processed through digestion is 1.44:1 (288,000 gallons per day:200,000 gallons per day). As was discussed in Section 5 of this report, operating at high primary sludge to total sludge mass ratios leads to digester stress conditions. It is recommended that for better overall process operation the primary to secondary flow rate volumes should be altered to 193,000 gallons per day:297,000 gallons per day (0.65:1).

6.1.6. Digester Gas Production

Calculations in Appendix M show that at the 490,000 gallon per day sludge processing rate under the conditions previously discussed, more gas will be generated than required to meet total sludge heating requirements. The average daily excess gas production is 658,000 cubic feet per day and will range from 780,000 cubic feet per day during the summer to 537,000 cubic feet per day during the winter.

6.2. EVALUATION OF THERMOPHILIC SYSTEM OPTION[2]

Review of thermophilic anaerobic digestion clearly indicates that the process should be seriously considered for the least cost, short term solution of the sludge processing and disposal problems of the District of Columbia's Wastewater Treatment Facility. The three significant advantages of the process (a) increased sludge processing capability, (b) improved sludge dewatering, and (c) increased destruction of pathogens, are all pertinent to the District's situation.

More detailed checks should be made on a number of items prior to deciding to convert the existing digesters to thermophilic operation. These include (a) amount and type of additional sludge heating required, (b) structural competency of existing digesters and piping at thermophilic temperatures, (c) needed improvements in the temperature control system, (d) equipment needed to remove the increased amounts of moisture to be expected from the digester gas, and (e) how to avoid possible inhibition by ammonia.

[2]All supporting calculations for Section 6.2 are given in Appendix N.

6.2.1. Ulitmate Disposal

Thermophilically digested sludge is generated under conditions that approach disinfection, thus allowing for acceptable final disposal to the land. As with the current mesophilic operation, it would be important to minimize transportation cost, therefore, the driest cake possible is desirable.

6.2.2. Digested Sludge Vacuum Filter Operation

Thermophilically digested sludge exhibits better filterability than straight mesophilically digested sludge--some times as much as double. In Appendix N, the analysis used a 25 percent improvement in yield resulting from thermophilic digestion, a value considered conservative. Calculations indicate that thermophilic digestion of all Blue Plains sludge would require the use of 8 of the 21 new vacuum filters for dewatering of the digested sludge.

Another advantage would be the elimination of the need for sludge conditioning by elutriation or by added iron salts. Thermophilically digested sludges can be conditioned using a combination of anionic and cationic polymers. This change alone would reduce the present amount of solids to be disposed of by 94 pounds per million gallon of influent flow--over 16 dry (96 wet) tons per day.

A disadvantage to dewatering this sludge is that the sludge must be cooled to under 90°F. Calculations in Appendix N indicate that the existing elutriation tanks might be used as cooling tanks. Potential odor problems may also be minimized by using the elutriation step because of this liquid to liquid cooling.

6.2.3. Impact of Increased Heat Requirements

The thermophilic digestion process being evaluated operates at 122°F. The existing heating capabilities at Blue Plains have been calculated to be inadequate, with approximately 16.1×10^6 BTU per hour additional heating needed during the winter months. It is suggested that direct steam injection be considered for supplying the additional heat since this has been used successfully in Los Angeles, Moscow, and the Canadian study. Structural competency should be checked by a structural engineer. A control engineer should be engaged to look at the temperature control system with a maximum variation of about ±1.5°F being permitted (at 120°F operation).

6.2.4. Digestion Tanks

Calculations indicate that the existing digestion tanks are capable of taking the full sludge load in the thermophilic range of operation if they are cleaned and grit accumulation kept to a minimum.

6.2.5. Volatile Matter Reduction and Gas Production

Analysis indicates that the percentage volatile matter reduction in the thermophilic system would be about the same as the mesophilic system but at one half the time. Calculations in Appendix N show that more gas will be generated than required to meet total sludge heating requirements. The average daily excess gas production is 840,000 cubic feet per day and will range from 1,000,000 cubic feet per day during the summer to 600,000 cubic feet per day during the winter.

6.2.6 Transition from Mesophilic to Thermophilic Operation

It would be desirable to make the transition from mesophilic to thermophilic operation as rapidly as possible. However, caution should be exercised in making this transition since very little information is available on the maximum rate at which this transition can be effected. In Garber's early work (4), he indicated that almost six months were needed to establish the first thermophilic unit as a separate culture. By seeding and more rapid increases in temperature, he was able to cut this time to three months.

In Garber's later work, he gave more details as to the transition procedure. This time he increased the temperature from 96°F to 126°F at the rate of 1°F per day while maintaining the load at approximately 0.1 pound volatile solids per cubic foot per day. This was not successful and the digester turned "sour." He then reduced the loading to a minimum while maintaining the temperature at the 126°F level but observed no change in condition over a four month period. The temperature was then reduced to 120°F and over a three week period, satisfactory digestion commenced.

Because of Garber' experience, personnel in Chicago were more cautious in raising the temperature. Their procedure was to raise the temperature at a rate of 1°F per day for only five days and then to wait for two to three weeks for the digester to stabilize before again increasing the temperature 1°F per day until a final temperature of 127°F was attained. During this period the loading on the digester was maintained at about 0.13 pounds volatile solids per cubic foot per day. Several surges (one up to 2,500 mg/liter) were observed during the transition period and it took approximately one year for the digester to stabilize at consistently low volatile acid concentrations.

From the above, it can be clearly seen that there is still much to be learned about the correct procedure for making the transition from mesophilic to thermophilic digestion. The three variables involved are (1) rate of change of temperature, (2) rate of change of loading, and (3) maximum temperature to be attained. Garber's experience indicates that, at least initially, the maximum temperature should be limited to 120°F.

In both the Los Angeles and Chicago experiences, the loading on the digester was maintained at its normal value while the temperature was gradually increased. An alternate approach would be to stop the loading to the digester, bring it to the new temperature as rapidly as possible, and then gradually increase the loading. Limited experience in Atlanta (12) during the summer of 1980 indicates that the transition might be accomplished more rapidly by following the latter procedure.

6.3. EVALUATION OF MESOPHILIC - THERMOPHILIC SYSTEM OPTION[3]

As noted at the beginning of Chapter 6, the primary purpose of this study was to evaluate the existing sludge processing operation with the intent of determining if a new, two-stage, mesophilic-thermophilic anaerobic digestion process could be applied to digest all sludge production using existing equipment and without substantial capital expenditures.

The mesophilic-thermophilic system offers all the advantages of thermophilic digestion (disinfection, increased volatile matter reduction, and improved dewatering) and produces an essentially innocuous sludge. In addition, the system is simple to operate and has built in buffering capacity to deal with unusual loading conditions. Disadvantages are that it requires heating of two completely separate digestion processes (one at 95°F, the other at 122°F) and that there is only one operational plant, handling 100,000 people, in the United States. Both the mesophilic-thermophilic and thermophilic processes will require that the sludge be cooled before being dewatered.

6.3.1. Ultimate Disposal

The sludge from this process is extremely inert and looks very similar to composted sludge. This quality should allow for disposal to both private and public lands.

6.3.2. Digested Sludge Vacuum Filter Operation

The mesophilic-thermophilic process will have the same vacuum filter operation requirements as for the thermophilic option.

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[3] All supporting calculations for Section 6.3 are given in Appendix O.

6.3.3. Impact of Increased and Dual Heat Requirements

The total sludge heating requirements for this process option are the same as the previous thermophilic option. The major difference is that the mesophilic digestion tanks are to be maintained at 95°F and the thermophilic tanks at 122°F. The suggested method of sludge heating is as follows:

- For the mesophilic tanks, the existing sludge heat exchangers would be used and steam (approximately 500,000 BTU per hour) would be injected into each of the tanks.

- For the thermophilic tanks, all heating would be by direct steam injection (approximately 5,000,000 BTU per hour per tank).

As was mentioned earlier, calculations indicate that the increased temperature will have no significant impact on sludge piping and wall structural integrity.

6.3.4. Digestion Tanks

Calculations indicate that even if all the existing tanks volume was available, the hydraulic detention time under each process condition would be inadequate. In order to process all of the currently generated sludge through this process, four more digesters would be required.

6.3.5. Volatile Matter Reduction and Gas Production

Limited experience indicates that a 50 percent volatile matter reduction can be expected from digesting the sludge. Calculations in Appendix O show that more gas will be generated than required to meet total sludge heating requirements. The average daily excess gas production is 1,109,000 cubic feet per day, and will range from 1,384,000 cubic feet per day during the summer to 833,000 cubic feet during the winter.

6.4. SUMMARY

Table 6-2 summarizes the major process considerations in implementing either of the three anaerobic digestion alternatives to fully process the maximum sludge quantities given in Table 6-1.

TABLE 6-2. SUMMARY OF MAJOR PROCESS CONSIDERATIONS IN IMPLEMENTING FULL ANAEROBIC DIGESTION OF BLUE PLAIN SLUDGES

	Mesophilic digestion of all sludge	Thermophilic digestion of all sludge	Mesophilic-thermophilic digestion of all sludge
Provisions for better grit removal	yes	no	yes
Digestion tanks required	18	12	16
Additional heat requirements BTU's x 10^6	4.4	16.1	16.1
Average daily revenues from excess digestion gas production, dollars[a]	1298	802	1057
Additional elutriation capacity	yes	no	no
Vacuum filters required to be operational	10	8	8
Maximum wet tons per day for disposal[b]	1148	1033	961

[a] Excess gas production sold to Naval R&D at $.95 per 1,000 cubic feet.

[b] Cake solids - 16.5 percent.

SECTION 7
ECONOMIC IMPACT

Municipal sludge management is a challenging field. It is also an expensive one. It is not unusual these days for a wastewater treatment facility to devote over fifty percent of its capital and operational budget to "managing" the steadily increasing amounts of wastewater sludge.

The Blue Plains Wastewater Treatment Plant is at a crossroads. The Plant's current wastewater flow generates almost 500,000 pounds per day of sludge. Several sludge treatment and disposal options are utilized:

- o Dewatering of thickened raw sludge followed by land trenching
- o Dewatering of thickened raw sludge followed by composting
- o Mesophilic anaerobic digestion of thickened raw sludges followed by dewatering and application to land.

In the near future, the land trenching alternative must be discontinued. Composting with its attendant problems and land requirements is not practical. The third existing disposal scheme--mesophilic anaerobic digestion--is a variable option.

This section of the report will not answer the question--what should be done? What it will do is show how economically attractive anaerobic digestion, in any of three process configurations, can be.

7.1. OPERATIONAL COST IMPACT

Table 7-1 presents average sludge generation per million gallons of influent flow for five different sludge management conditions.

CONDITION I - This is historical information and represents the situation for the first six months of 1980. During this time period, approximately 40 percent of the sludge was digested, 40 to 45 percent went to land trenching, and 15 to 20 percent to composting.

CONDITION II - This is a projection of what the situation will be in late 1981 based on completion of current on-going process changes. At this time, approximately 40 percent of the sludge will be digested, the rest will be composted.

CONDITION III- This is a projection of what the situation would be if all the sludge was processed through a mesophilic anaerobic digestion system.

CONDITION IV - This is a projection of what the situation would be if all the sludge was processed through a thermophilic anaerobic digestion system.

CONDITION V - This is a projection of what the situation would be if all the sludge was processed through a proposed new process modification, mesophilic-thermophilic anaerobic digestion.

The economic impact of the numbers shown in Table 7-1 are better illustrated in Figures 7-1 and 7-2.

In Figure 7-1 a comparison is made of the average cost for chemicals and final disposal per million gallons of influent flow for the five conditions given in Table 7-1. The 40 percent reduction in cost shown between the existing 1980 and projected 1981 plan is solely due to eliminating land trenching and replacing it with composting. The significant cost reductions shown between the projected 1981 plan and the three anaerobic diges-

TABLE 7-1. SUMMARY OF AVERAGE SLUDGE GENERATION PER MILLION GALLONS OF INFLUENT FLOW

	Based on January to June 1980		Based on completion of current on-going changes planning-fall 1981		Based on mesophilic digestion of all sludge		Based on thermophilic digestion of all sludge		Based on mesophilic-thermophilic digestion of all sludge	
	Sludge generated per MGIF,[a] pounds	Cost per MGIF, dollars	Sludge generated per MGIF, pounds	Cost per MGIF, dollars	Sludge generated per MGIF, pounds	Cost per MGIF, dollars	Sludge generated per MGIF, pounds	Cost per MGIF, dollars	Sludge generated per MGIF, pounds	Cost per MGIF, dollars
Primary sludge	696		696		696		696		696	
Secondary sludge	568		656		656		656		656	
Gravity thickening	0		0		0		0		0	
Dissolved air flotation thickening	2	3.83[b]	3	4.42	3	4.42	3	4.42	3	4.42
Anaerobic digestion	(512)	(5.84)[c]	(510)	(5.81)	(510)	(5.81)	(510)	(5.81)	(572)	(6.52)
Elutriation	0		0		0		0		0	
Dewatering digested sludge										
° Iron salt	86	5.89[d]	98	6.71	98	6.71				
° Polymer							2	8.00	2	8.00
Dewatered raw sludge										
° Iron salt	98	6.71	88	6.03						
° Lime	446	37.86[e]	238	20.26						
Disposal digested sludge		38.18[f]		42.86		42.86		38.50		35.54
Disposal raw sludge		147.32[g]		65.36						

[a] MGIF = million gallons of influent flow.

[b] Polymer cost averaged $13.50 per ton of dry feed solids.

[c] Based on 12 cubic feet per pound volatile matter reduced and $0.95 per 1000 cubic feet.

[d] Based on $0.0685 per pound of this sludge produced.

[e] Based on an average cost of $0.0849 per pound of this sludge produced.

[f] Based on a $15 per wet ton disposal cost.

[g] Based on a $35 per wet ton composite cost (some composting, some trenching).

[h] Based on a $24 per wet ton composting cost.

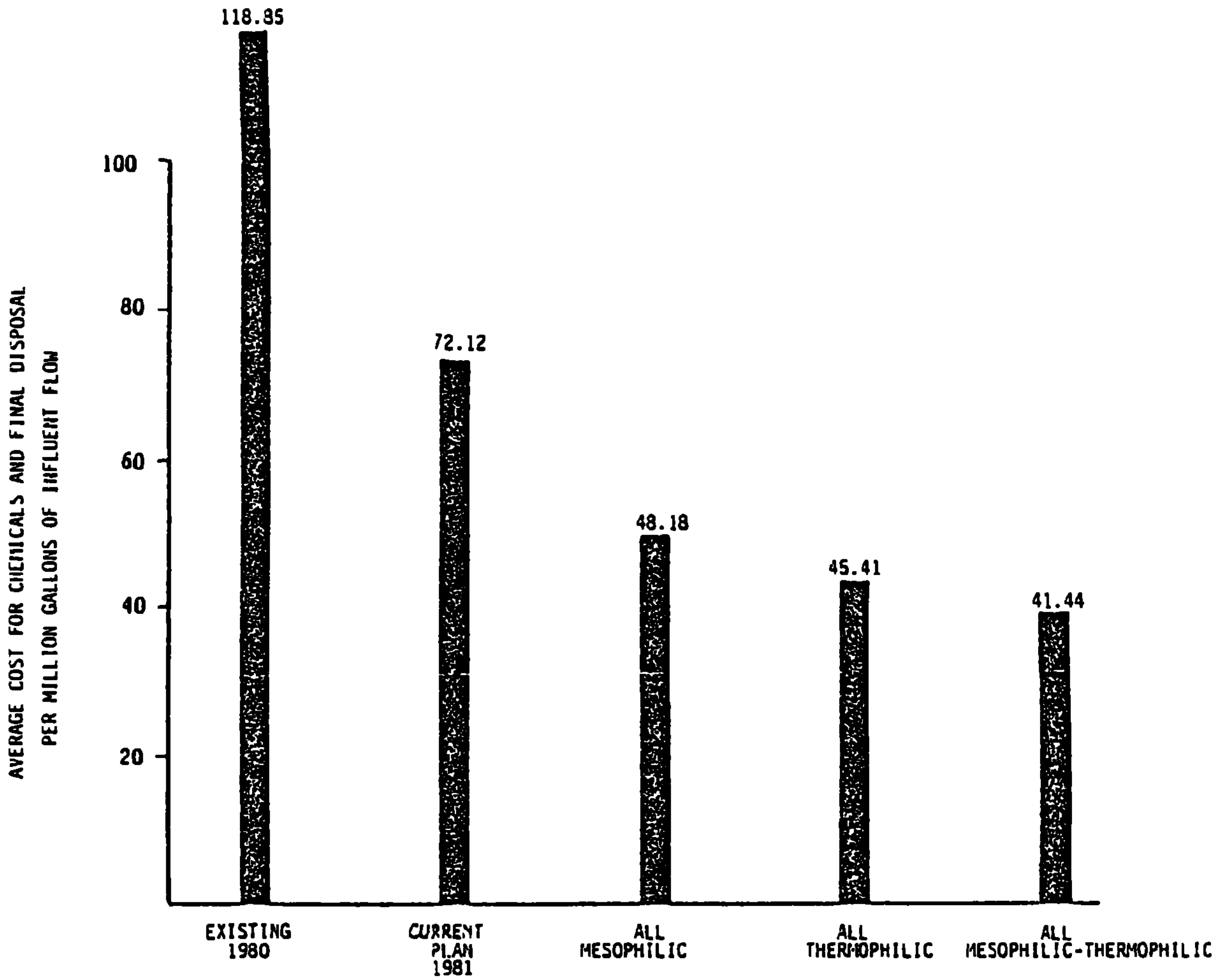

Figure 7-1. Average cost for chemicals and final disposal per million gallons of influent flow for the five conditions given in table 7-1.

tion alternatives are due to two items. First, in digestion approximately 37 to 40 percent of the raw solids generated are destroyed so that digestion of all the sludge will significantly reduce the mass of solids that requires disposal. Secondly, the dewatering operation for digested sludge at Blue Plains does not require lime for conditioning: in the 1981 projection, the raw sludge dewatering operation will require a significant amount of lime, which will increase solids by 238 pounds per million gallons of influent flow. The higher unit cost of mesophilic over the other two digestion alternatives is mainly because the other two digestion alternatives do not require iron salts in the dewatering operation (equivalent to 96 pounds of solids per million gallons of influent flow). Meso-thermophilic digestion has a lower overall cost than thermophilic digestion because of a higher volatile matter reduction.

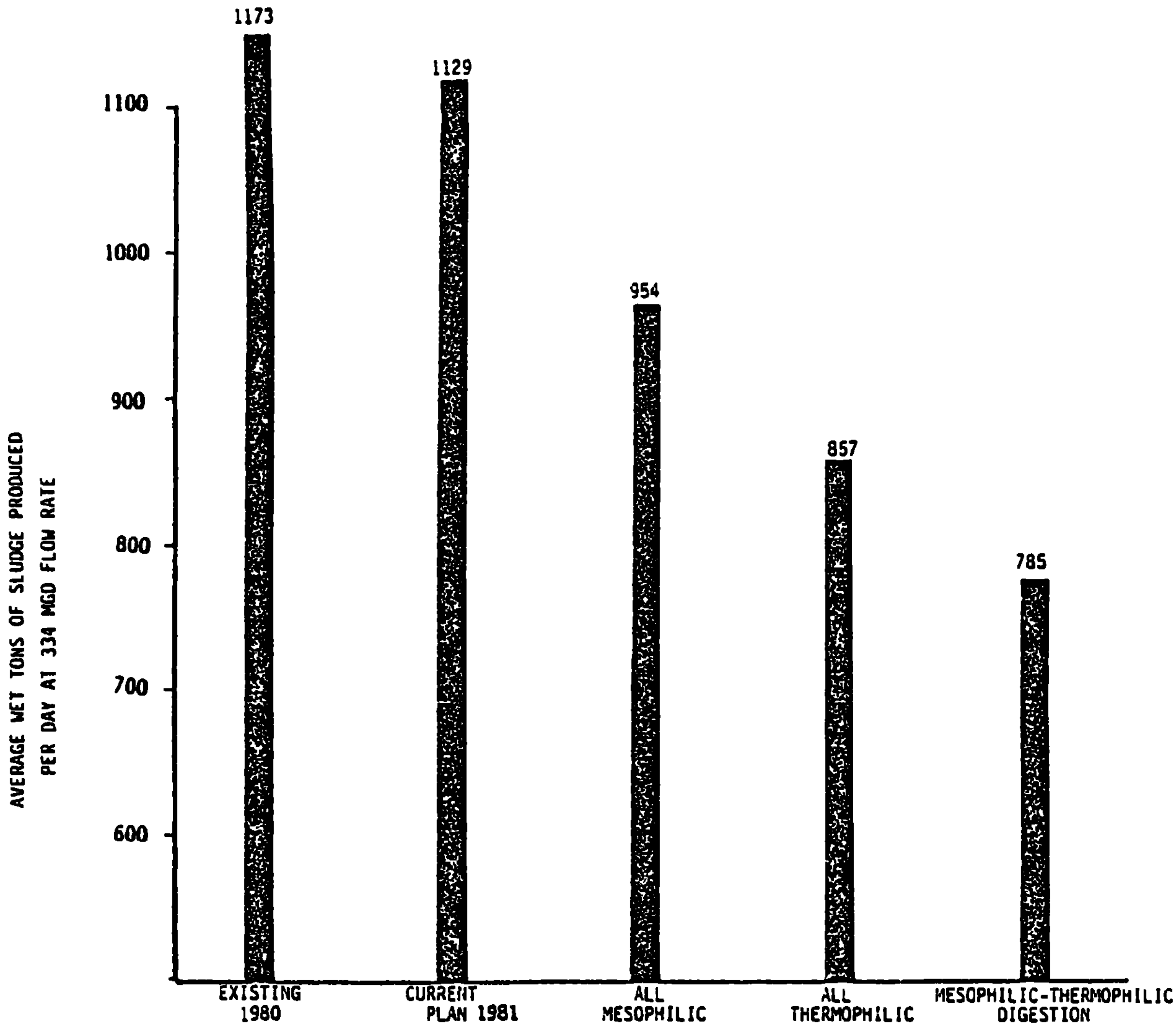

Figure 7-2. Average wet tons of sludge produced per day for the five conditions given in table 7-1.

Figure 7-2 shows the difference in wet tons generated by the five different conditions given in Table 7-1. The large reduction in wet tons is due to two items. First, in the digestion alternatives approximately 37 to 40 percent of the raw solids are destroyed (converted into methane gas, carbon dioxide, and water); secondly, significantly less chemicals are required in the dewatering operations.[1] The difference in wet tons between the three digestion alternatives is due to lower chemical conditioning requirements for the thermophilic and meso-thermophilic options, and the higher volatile matter reduction achieved by the meso-thermophilic option.

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[1] For every million gallons of flow into the plant, the raw sludge dewatering option must dispose of 738 more pounds of solids. This is equivalent to 3432 wet pounds or 1.72 wet tons.

Of the three anaerobic digestion alternatives, the mesophilic-thermophilic option is shown to have the lowest operational cost. In addition, of the three digestion alternatives only the mesophilic-thermophilic option produces a sludge that is extremely inert, with essentially no pathogens and little to no odor potential. These benefits should allow the District of Columbia to develop a strong marketing campaign to allow disposal of dewatered digested sludge on public and private lands.

7.2. CAPITAL COST IMPACT

In order for any of the three anaerobic systems to process all the sludge currently being generated at Blue Plains, some capital expenditures would be required to expand and upgrade existing equipment. A detailed cost analysis was not performed. However, the following unit operation analysis indicates that the capital improvement cost would be the greatest for the mesophilic system and the least for the thermophilic system. The mesophilic-thermophilic system would lie between the two.

7.2.1. Improvements in Grit Removal

The analyses in Sections 4 and 5 indicated that 1.4 to 2.0 cubic feet of grit per million gallons of influent flow is contained in the sludge stream. Grit is inert, but it reduces the anaerobic digestion process capacity by occupying tank volume. The calculations of tank volume (in Section 6) assume minimal grit accumulation, but at the current time approximately 25 to 30 percent of the existing digestion tank capacity is occupied by grit. The best options are to increase grit removal before the influent flow reaches the digesters or to increase digestor capacity to allow for accumulation. The capital expenditures for grit removal equipment are approximately equal for all digestion processes.

The cost for the necessary grit removal equipment has not been estimated in this study, but it should be approximately equal for all three digestion options. However, if grit removal is not provided, then the digestion capacity would have to be increased by about 25% to allow for accumulation.

7.2.2. Digestion Tanks

Assuming that grit is removed upstream of the digestors, the need for additional digestion tanks to handle the entire sludge stream is approximately six for mesophilic, four for the mesophilic-thermophilic, and none for the thermophilic process. A preliminary analysis indicated that operating the digesters at the thermophilic temperature would not cause structural problems.

7.2.3. Sludge Heating Requirements

All three processes would require additional heating capacity to handle the entire sludge stream. The mesophilic option may be satisfied by existing in-house heat generation. The other two options would require auxiliary steam generating equipment to allow direct steam injection into each digestion tank operating at thermophilic temperatures.

7.2.4. Improvements in Gas Piping

The existing gas collection piping and safety devices need to be replaced or upgraded for any of the anaerobic digestion processes. The piping shows corrosion damage in places, and some of the gas protection equipment is not functional.

7.2.5. Improvements in Mixing

The existing digestor mixing is believed to be deficient for several reasons. First, the majority of the existing gas-mixing equipment was originally installed in 1960 on a temporary basis and was not designed as a complete tank mixing system. Secondly, the existing internal heating coils allow build up of grit and inert material not capable of being mixed. Improvements in grit removal, removal of internal heating coils, and an upgrade of the gas-mixing system to current standards would significantly improve the mixing operation and, hence, improve the tank utilization.

7.2.6. Elutriation Capacity

The mesophilic system would require a 200 percent increase in elutriation tank capacity. The mesophilic-thermophilic and thermophilic processes both could not require elutriation.

7.3. SUMMARIZED ECONOMIC IMPACT

The operating cost analysis has developed a sludge handling unit cost for the five options considered in this study. The developed costs are not necessarily "accurate" (in the sense that all minor cost factors have been included), but they are consistent and reflect the comparative costs of the options. The unit costs are presented in Table 7-2, which also includes a comparative presentation of capital costs for the five options.

A qualitative analysis of the disposal factors also is summarized in Table 7-2. This analysis is reflected to some extent in the disposal cost analysis for the current and projected 1981 options, in that disposal costs of undigested sludge are included. The anaerobic digestion options for the entire sludge stream will result in a reduced disposal cost. The characteristics of the sludge product as indicated in the disposal factors in Table 7-2 will determine the eventual disposal cost.

TABLE 7-2. SUMMARY OF OPERATING AND CAPITAL COST REQUIREMENTS FOR SLUDGE HANDLING OPTIONS

Option[a]	Operating cost[b] ($/MGIF)	Capital cost items[c]					
		Grit removal	Digestion tanks	Sludge heating	Gas piping	Mixing	Elutriation
1. Current (January to June 1980)	118.85	NA	NA	NA	NA	NA	NA
2. Projected 1981	72.12	NA	NA	NA	NA	NA	NA
3. Mesophilic, all sludge	48.18	Yes, even	Yes	Yes	Yes	Yes	Yes
4. Mesophilie-thermophilic, all sludge	41.44	Yes, even	Yes	Yes, +	Yes	Yes	No
5. Thermophilic, all sludge	45.41	No	No	Yes, +	Yes	Yes	No

[a]Current and projected options do not include the digestion of all sludge generated.

[b]MGIF = million gallons of influent flow.

[c]Yes = a capital expenditure is required or recommended to handle the entire sludge shown.

No = existing equipment is adequate or not required to treat the entire sludge stream

NA = not applicable to the evaluation.

Even = approximately the same capital would be required

Number = the number of digestors required.

+ = a significantly greater capital expense than for the other options is expected.

REFERENCES

U.S. EPA 1979. Process design manual: sludge treatment and disposal. EPA-625/1-79-001. U.S. Environmental Protection Agency, Cincinnati, Ohio.

APPENDIX A

JANUARY TO JUNE 1980
SUMMARY OF AVERAGE MONTHLY OPERATING
DATA ON BLUE PLAINS SLUDGE MANAGEMENT OPERATIONS

The data shown in Appendix A were obtained from operating log sheets and from conversations held with Mr. Ed Jones, Mr. Steve Bennett, and Mr. Walt Baily.

The purpose of this analysis was to perform solids mass balances around each unit process and to segregate the type of solids--primary, waste activated, lime, ferric chloride--involved in each situation. This is possible at Blue Plains since extensive process stream testing is conducted on a regular basis. Analysis of six months of operating data indicates the following:

1. On a total solids basis, mass balance calculations around each unit process are generally within 5 to 10 percent of closing.

2. Existing flow meters should be maintained and calibrated on a regular basis. Many of the existing flow meters have been out of service for some time. Flow calculations are approximated using either pump strokes, pump curves, or changes in tank liquid levels. This type of hydraulic data makes for difficult process control and is believed to be responsible for some of the difficulty in closing the mass balances around each process.

3. Numerous duplication of data on various log sheets is currently required. In many cases different data are indicated for situations in which the data should be identical. It is recommended that a review of all existing log sheets be conducted, that they be consolidated where possible, and that better quality control be maintained over the logging of data.

4. Some of the solids analyses are done by total solids and some by total suspended solids. Since total solids is equivalent to dissolved solids plus suspended solids, a difficulty exists in doing solids mass balances.

   This difficulty arises for several reasons:

   - Dissolved solids do not concentrate, but do increase in concentration as the sludge moves through the sludge treatment process.

   - Analyses of thickened or dewatered sludges show that the amount of dissolved solids is almost insignificant compared to the suspended solids.

- Analyses of clarified liquors, filtrates, or elutriates show that suspended solids normally comprise only 15 to 50 percent of the total solids, while dissolved solids normally comprise the majority.

TABLE A-1. SUMMARY OF RECENT AVERAGE MONTHLY HISTORICAL DATA ON RAW SLUDGE QUANTITIES[a]

		Primary sludge					Waste activated sludge				
			Total dry solids[b]		Total dry volatile solids			Total dry solids[c]		Total dry volatile solids	
Year	Month	Volume, MGD	Percent	Tons per day[d]	Percent volatile	Tons per day	Volume, MGD	Percent dry solids	Tons per day[e]	Percent volatile	Tons per day
1980	Jan	4.2	0.55	98	77	75	2.7	0.91	103	66	68
	Feb	3.8	0.72	116	81	94	3.1	0.72	94	68	63
	Mar	3.7	0.61	96	78	75	2.9	0.67	82	66	54
	Apr	4.4	0.78	146	78	114	3.4	0.73	104	66	69
	May	4.5	0.62	119	79	94	2.8	0.68	80	62	65
	Jun	3.0	0.92	117	85	99	2.7	0.88	101	64	65

[a] Data obtained from Mr Walt Baily and Blue Plains operating records.

[b] There are strong indications that significant amounts of grit are in the primary sludge. Detailed analysis of volume was not made, but currently is estimated at 300 to 350 cubic feet per day or approximately 30,000 to 35,000 pounds per day

[c] Iron is added for phosphorus removal. The iron - phosphate sludge contribution is between 11 to 15.5 percent of the total mass

[d] Density of this material is 8.5 pounds per gallon

[e] Density of this material is 8.39 pounds per gallon.

TABLE A-2. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON GRAVITY THICKENING[a]

		Total gravity thickener inflow[b]					Gravity thickener underflow				
			Total dry solids[c]		Total dry volatile solids			Total dry solids		Total dry volatile solids	
Year	Month	Volume, MGD	mg/l	Tons per day	Percent volatile	Tons per day	Volume, MGD	Percent	Tons per day[d]	Percent volatile	Tons per day
1980	Jan	16 0	1600	107	74.7	74 7	328	7.3	105	79	74
	Feb	15.9	1915	127	67 4	82.2	437	6.7	129	75	97
	Mar	15.7	1600	105	69.4	70.8	.365	6.7	108	70	76
	Apr	15 7	2350	154	66 5	99.1	.521	6 9	158	65	103
	May	15.3	1960	125	72.0	83.5	.435	6 4	120	66	79
	Jun	16.0	1845	123	69.7	82.8	370	7.9	129	71	92
	Jul	16.9			67.6						

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data. Data are not available on gravity thickener overflow

[b] Total gravity thickener inflow consists of flow from primary clarifiers, dilution water, approximately 50 percent of the elutriation tank overflow

[c] Existing data are considered to be incorrect because they indicate a sludge mass greater than the output being disposed of At this time, the numbers shown for tons per day are based on primary sludge production (Table A-1) + the suspended solids contribution of the dilution water (0.17 to 1 0 tons per day) + one half of the total solids in the elutriate overflow (Table A-5)

[d] Density of this material is 8.8 pounds per gallon.

TABLE A-3. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON DISSOLVED AIR FLOTATION THICKENING[a]

		Total dissolved air flotation thickener inflow[b]					Thickened material				
			Total dry solids		Total dry volatile solids			Total dry solids		Total dry volatile solids	
Year	Month	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day
1980	Jan	2 7	0 91	103	66.0	68	0 665	3.7	102 7	66 7	68 5
	Feb	3 1	0 72	93	67 7	63	0.570	3.9	92 7	66 5	61 6
	Mar	2 9	0 70	85	65 9	56	0.496	4.1	84.8	69 2	58 7
	Apr	3 4	0.73	103	66.0	68	0.582	4.2	101 9	66 9	68 2
	May	2 8	0 68	79	62 0	49	0.448	4 2	78 4	63.8	50 0
	Jun	2.7	0.89	100	64.0	64	0.508	4.7	99 6	63 7	61 4

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data.

[b] Current polymer usage cost 12 to 15 dollars per ton of dry feed solids to dissolved air flotation thickeners.

(continued)

TABLE A-3. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON DISSOLVED AIR FLOTATION THICKENING[a] (continued)

Year	Month	Subnatant Flow: Volume, MGD	Subnatant Flow: Total dry suspended solids, mg/l	Subnatant Flow: Total dry suspended solids, Tons per day
1980	Jan	2.6	28	0.3
	Feb	2.9	21	0.3
	Mar	2.2	21	0.3
	Apr	7.7	37	1.2
	May	3.1	44	0.6
	Jun	5.5	17	0.4

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data.

TABLE A-4. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON ANAEROBIC DIGESTION[a]

		Primary sludge					Waste activated sludge				
			Total dry solids		Total dry volatile solids			Total dry solids		Total dry volatile solids	
Year	Month	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day
1979	Jan	0.242	8 0	80 7	67.0	54 1	0.174	5.9	42.8	68 7	29 4
	Feb	0.188	8 6	67 4	73.7	49.7	0.170	5.2	36 9	67.7	25 0
	Mar	0.284	8 3	98.3	73 1	71.8	0.100	5.6	23 3	66 0	15 4
	Apr	0 301	7.3	91.6	75.0	68.7	0 162	4.5	30 4	67.1	20 4
	May	0.199	9.8	81 3	72.1	58 6	0.243	5.4	54 7	64.8	35 4
	Jun	0 225	8 8	82.6	70.4	58.1	0.206	5.2	44.7	63 3	28 3
	Jul	0 194	7 3	59 0	66.1	39.0	0 229	4.9	46 8	61.6	28 8
	Aug	0.216	7 8	70.3	64.6	45 4	0.239	5.1	50 8	63 5	32 3
	Sept	0 177	7 4	54 6	61.3	33.5	0.283	5.0	59.0	62.9	37 1
	Oct	0 166	7 2	49.8	64 1	31 9	0 275	4.6	52 8	66 5	35 1
	Nov	0 177	---	----	----	----	0.141	---	----	----	----
	Dec	0 195	6.4	52.0	72 5	37 7	0 162	4.1	27 7	66 7	18 5
1980	Jan	0.228	7 3	69.4	78.5	54 5	0.136	3.7	21 0	66 5	14 0
	Feb	0.227	6.7	63 4	75 0	47 6	0 276	3 9	44 9	69.2	31 1
	Mar	0.228	6 7	63.7	69.6	44.3	0.293	4.1	50.1	66 9	33.5
	Apr	0.196	6 9	56.4	65.4	36.9	0.269	4.2	47.1	63.8	30 6
	May	0 222	6 4	59 2	66.3	39.2	0.242	4 2	42.4	63.7	27 0
	Jun	0.169	7 9	55.7	71 0	39.5	0.316	4.7	62.0	66 3	41 1

[a] Data obtained from Mr Steve Bennett and Blue Plains operating data.

(continued)

TABLE A-4. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON ANAEROBIC DIGESTION[a] (continued)

		Combined digester input						
			Total dry solids		Total dry volatile solids		Waste activated sludge fraction	
Year	Month	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day	Percent by weight	Percent by volume
1979	Jan	0 416	7.1	123.5	67 6	83.5	35	42
	Feb	0 358	7.0	104.3	71 6	74 7	35	48
	Mar	0.384	7.6	121.6	71.7	87.2	19	26
	Apr	0 463	6.3	122.0	73.0	89 1	25	35
	May	0.442	7 4	136.0	69.1	94.0	40	55
	Jun	0 431	7 1	127.3	67 9	86.4	35	48
	Jul	0 423	6.0	105.8	64.1	67 8	44	54
	Aug	0 455	6.4	121.1	64 2	77 7	42	53
	Sept	0.460	5.9	113.6	62.1	70.6	52	62
	Oct	0.441	5 6	102 6	65.3	67.0	51	62
	Nov	0.318	---	-----	----	----	--	44
	Dec	0.357	5.4	79.7	70.5	56.2	35	45
1980	Jan	0.364	5.9	90.4	75.8	68.5	23	37
	Feb	0 503	5.2	108.3	72.6	78.7	41	55
	Mar	0 521	5 2	113.8	68.4	77.8	44	56
	Apr	0.465	5.3	103.5	65.2	67.5	46	58
	May	0.464	5.2	101 6	65.1	66.2	42	52
	Jun	0 485	5.8	117.7	68.5	80.6	53	65

[a] Data obtained from Mr Steve Bennett and Blue Plains operating data

(continued)

TABLE A-4. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON ANAEROBIC DIGESTION[a] (continued)

		Digested sludge									Digester gas		
			Total dry solids		Total dry volatile solids								
Year	Month	Volume, MGD	Percent dry solids	Tons per day	Percent volatile	Tons per day	Temp.,[b] °F	Volatile acids, mg/l	Alk.[c] mg/l	NH_3-N mg/l	Gas produced,[d] cubic feet x1000	CO_2 content, percent	Gas produced per pound of volatile solids reduced
1979	Jan	0.416	3.3	57.2	54.2	31.0	95	168	5629	1107	912	32.7	8.7
	Feb	0.358	3.3	49.3	54.6	26.9	96	233	6229	1451	846	33.4	8.8
	Mar	0.384	3.4	54.4	54.2	29.5	94	204	5956	1428	834	34.4	7.2
	Apr	0.463	2.9	56.0	55.7	31.2	95	158	5313	1374	999	31.8	8.6
	May	0.442	3.2	59.0	55.5	32.7	96	172	5868	1400	1216	32.6	9.9
	Jun	0.431	3.8	68.3	53.3	36.4	95	260	6488	1568	1205	32.6	12.0
	Jul	0.423	3.8	67.0	55.3	37.1	95	207	6578	1302	886	32.8	14.4
	Aug	0.455	3.6	68.3	54.3	37.1	95	183	5742		749	29.1	9.2
	Sept	0.460	3.9	74.8	52.3	39.1	94	176	5238		797	33.0	12.6
	Oct	0.441	3.6	66.2	53.8	35.6	95	166	5209		849	31.9	13.5
	Nov	0.318	3.6	47.7	56.5	27.0	95	180	5099		808	34.9	----
	Dec	0.357	2.7	40.2	56.6	22.7	94	225	7272		879	35.0	9.6
1980	Jan	0.364	2.3	34.9	57.6	20.1	92	142	6784		1061	35.5	11.0
	Feb	0.503	2.6	54.5	55.3	30.2	91	126	7624		1246	34.3	12.8
	Mar	0.521	2.6	56.5	55.1	31.1	93	284	8403		1123	35.1	12.0
	Apr	0.465	2.8	54.3	56.1	30.5	90	338	8914		888	34.6	12.0
	May	0.464	3.1	60.0	55.5	33.3	93	371	8328		906	34.2	13.8
	Jun	0.485	3.1	62.5	54.8	34.2	95	323	8366		994	34.9	10.7
	Jul		3.1		55.3		93	329	8140		928	34.8	

[a] Data obtained from Steve Bennett and Blue Plains operating data.

[b] Temperature measured on recycle sludge before entering heat exchanger.

[c] AlK - alkalinity. Alkalinity is measured on entire solids mass rather than on subnatant of sludge.

[d] Blue Plains is in the process of signing a contract with Naval Research to sell all excess digester gas for $0.95 per 1000 cubic feet.

TABLE A-5. SUMMARY OF RECENT AVERAGE MONTHLY HISTORICAL DATA ON ELUTRIATION TANK OPERATIONS[a]

		Anaerobic digesters					Wash water[b]		
			Total dry solids		Total dry volatile solids			Total dry suspended solids[c,d]	
Year	Month	Volume, MGD	Percent dry solids	Tons per day[e]	Percent Volatile	Tons per day	Volume, MGD	mg/l	Tons per day
1980	Jan	0.397	2.6	45	57	25.3	2.27	18	0 17
	Feb	0.509	2.4	53	57	30.5	2.34	12	0 12
	Mar	0 509	2.4	53	56	29 7	2.32	27	0.26
	Apr	0.466	2 7	54	52	28.2	2.28	13	0 12
	May	0.448	3 0	58	52	30.4	2 17	20	0 18
	Jun	0 492	3 3	70	54	38.0	2 21	9	0 08

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating records.

[b] Approximately half of the wash water is pumped from the River and the other half is secondary effluent before nitrification.

[c] Data indicate that solids contributed by washwater are insignificant compared to total mass through elutriation system

[d] Data indicate that solids contributed by polymer addition are insignificant compared to total mass through elutriation system.

[e] Density of this material is 8 64 pounds per gallon.

(continued)

TABLE A-5. SUMMARY OF RECENT AVERAGE MONTHLY HISTORICAL DATA ON ELUTRIATION TANK OPERATIONS[a] (continued)

		Elutriation underflow to dewatering			Elutriation overflow[b]		
			Total dry solids			Total dry solids[c]	
Year	Month	Volume, MGD	Percent dry solids	Tons per day[d]	Volume, MGD	mg/l	Tons per day
1980	Jan	0.227	5.4	54	2.44	1800	18.3
	Feb	0.279	4.8	59	2.57	1974	21.1
	Mar	0.309	3.6	49	2.52	1652	17.4
	Apr	0.187	5.5	46	2.56	1517	16.2
	May	0.207	5.4	52	2.41	1254	12.6
	Jun	0.241	5.2	55	2.46	1208	12.4

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating records.

[b] Elutriation overflow normally is supposed to return to primary treatment by gravity flow. At the present time, only part of the overflow flows by gravity to primary treatment The other part is pumped to gravity thickening. There are no flow meters on either line, and it is not possible to tell how much flow is going to gravity thickening or to primary treatment.

[c] Suspended solids for the same time period in mg/l were: Jan-295; Feb-874; Mar-1048; Apr-684; May-479; June-559.

[d] Density of this material is 8.85 pounds per gallon.

TABLE A-6. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON RAW SLUDGE DEWATERING[a]

		Primary sludge from gravity thickener			Waste activated sludge from dissolved air flotation thickeners			Lime added as dry CaO		Ferric chloride added as dry $FeCl_2$	
			Total dry solids			Total dry solids,					
Year	Month	Volume, MGD	Percent dry solids	Tons per day[b]	Volume, MGD	Percent dry solids	Tons per day[c]	Tons per day[d]	Dollars per day[e]	Tons per day	Dollars per day[f]
1980	Jan	0.100	7.3	32	0.529	3.7	82	55	3489	20.5	1488
	Feb	0.210	6.7	62	0.294	3.9	48	48	3045	17.8	1292
	Mar	0.137	6.7	40	0.203	4.1	35	38	2410	11.5	835
	Apr	0.325	6.9	99	0.313	4.2	55	48	3045	11.5	835
	May	0.213	6.4	60	0.206	4.2	36	47	2981	11.5	835
	Jun	0.201	7.9	70	0.192	4.7	38	48	3045	12.5	908

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data.

[b] Density of this material is 8.8 pounds per gallon.

[c] Density of this material is 8.39 pounds per gallon

[d] Blue Plains uses high grade lime, CaO, with 5 percent inerts. Value indicated is dry CaO and does not include 5 percent inerts.

[e] Summer of 1980-Blue Plains pays $0.031215 per pound CaO during the week and $0.034715 per pound CaO on holidays and Sundays.

[f] Summer of 1980-Blue Plains pays $0.0825 per pound of iron (Fe), which comprises 44 percent of ferric chloride ($FeCl_2$).

(continued)

TABLE A-6. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON RAW SLUDGE DEWATERING[a] (continued)

		Dewatered raw sludge off vacuum filter					Vacuum filter filtrate[b]			Washwater[c]		
			Total dry solids		Total dry volatile solids			Total dry solids			Total dry solids	
Year	Month	Wet tons per day[d]	Percent dry solids	Tons per day[d]	Percent volatile	Tons per day	Volume, MGD	mg/l[e]	Tons per day[f]	Volume, MGD	mg/l	Tons per day
1979	Aug		19 6		44 6							
	Sept		20 7		43.3							
	Oct		19.9		44 8							
	Nov		----		----							
	Dec		17.6		48.1							
1980	Jan	948	17.2	163	46.2	75	0.499					
	Feb	849	17 9	152	48.8	74	-----					
	Mar	490	21.6	106	41.0	44	-----					
	Apr	910	21.1	192	40.9	78	0.574					
	May	635	21 1	134	40 2	54	-----					
	Jun	695	22.6	157	42 1	66	0 440					
	Jul		22 1		41 4							

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data.

[b] Data base is extremely limited.

[c] Washwater volume is significant but no data are taken on this stream Suspended solids are not believed to be significant

[d] See footnote c on Table A-8.

[e] Analysis based on several samples indicated that total solids were about 500 mg/l and that suspended solids about 250 mg/l.

[f] Based on limited data, it is estimated that the total solids in the filtiate average 2 tons per day and can be considered insignificant.

TABLE A-7. SUMMARY OF RECENT AVERAGE MONTHLY DATA ON DIGESTED SLUDGE DEWATERING[a]

		From elutriation			Dry ferric chloride added		Dewatered digested sludge to disposal					Vacuum filter filtrate[b]		
			Total dry solids,					Total dry solids to disposal		Total dry volatile solids To disposal			Total dry solids	
Year	Month	Volume, MGD	Percent Dry Solids	Tons per day[c]	Tons per day	Dollars per day[d]	Wet tons per day	Percent dry solids	Tons per day	Percent volatile	Tons per day	Volume, MGD	mg/l[e]	Tons per day
1979	Aug						368	17.8	65.5	49.5	32.4			
	Sept						361	17.9	64.6	47.7	30.8			
	Oct						345	16.3	56.2	49 6	27 9			
	Nov						290	----	----	----	----			
	Dec						289	16 1	46.5	54.3	25 0			
1980	Jan	0 227	5 4	54	10.2	740	293	16.0	49.9	56.3	26 4	0.136		
	Feb	0.279	4 8	59	11.9	864	395	15 8	62.4	57.2	35.7	0 205	9600	8.2
	Mar	0 309	3 6	49	10.8	784	337	15.6	52.6	54.1	28.4	0.192	8800	7.0
	Apr	0.187	5.5	46	10.0	726	302	16.7	50.4	51.6	26.0	0.130	12000	6.5
	May	0 207	5.4	52	10.7	777	314	16.8	52.7	51.9	27.4	0.144	9200	5.5
	Jun	0.241	5.2	55	10.3	748	328	16.9	55.4	52.5	29.1	0.175	8200	6 0
	Jul						294	16.8	49.4	53.8	26.6			

[a] Data obtained from Mr. Steve Bennett and Blue Plains operating data.

[b] Data indicate that solids contributed by filtrate are insignificant compared to total mass of solids included.

[c] Density of this material is 8.85 pounds per gallon.

[d] Summer of 1980 Blue Plains pays \$0.0825 per pound of iron (Fe), which comprises 44 percent of ferric chloride ($Fe\ Cl_2$).

[e] Suspended solids for this same time period in mg/l were: Jan-185; Feb-66; Mar-110; Apr-103; June-270.

TABLE A-8. SUMMARY OF RECENT AVERAGE MONTHLY HISTORICAL DATA ON RAW SLUDGE DISPOSAL[a]

Year	Month	Raw sludge hauled from plant,[b,c] Wet tons per day	Trucking and disposal cost,[d] Dollars per day	Total dry solids, to disposal: Percent dry solids as trucked[e]	Total dry solids, to disposal: Tons per day	Break down of total dry solids to disposal: Raw sludge solids, tons per day	Break down of total dry solids to disposal: From lime addition tons per day[f]	Break down of total dry solids to disposal: From ferric chloride addition tons per day[g]
1980	Jan	1153	41,175	14.1	163	112	41	11
	Feb	1047	37,200	14.5	152	106	36	10
	Mar	758	26,362	14.0	106	72	28	6
	Apr	1004	35,587	19.1	192	150	36	6
	May	862	30,262	15.5	134	93	35	6
	Jun	762	26,512	20.6	157	104	36	17
	Jul	716	24,787					

[a] Data obtained from Mr. Ed Jones, Mr. Steve Bennett, and Blue Plains operating records.

[b] At the time of this report - Sept. 1980 - dewatered raw sludge was being disposed of as follows:

- Approximately 100 wet tons per day to Beltsville at an approximate trucking cost of $30/wet ton.
- Between 150 to 200 wet tons per day to MTI for composting at an approximate trucking cost of $30/wet ton. Contract on a year to year basis. Could possibly go up to 400 wet tons per day in the future.
- Balance of sludge to trenching with hauling and disposal cost at $35 to $40/wet ton.

[c] Wet tons of raw sludge hauled from plant are greater than wet tons per day of raw sludge produced by vacuum filters (Table A-6). Analysis of the dewatered cake dryness is done on the dewatered material as it discharges off the vacuum filter. Dewatered cake falls onto a conveyor and is conveyed to a sludge storage bin and then discharged to the haul trucks. Because of the design of the bin, the dewatered sludge does not discharge properly and water is used to move the material. The exact amount of water used is not known but is estimated between 3600 to 5500 gallons per day (15 to 23 tons per day).

[d] Based on 100 wet tons/day to Beltsville at $30/wet ton, 175 wet tons/day to MTI at $30/wet ton, and balance to trenching at $37.50/wet ton.

[e] Percentage calculated by taking ratio of tons per day dry solids divided by wet tons per day of sludge hauled from plant and multiplying by 100.

[f] Blue Plains uses high grade lime, CaO, with 5 percent inerts. Value indicated includes inerts and assumes that all Ca added leaves as a solid with the sludge.

[g] It is assumed that all iron (Fe) solids are in the sludge and that 20 percent of the chloride is in the sludge.

TABLE A-9 SUMMARY OF RECENT AVERAGE MONTHLY HISTORICAL DATA ON DIGESTED SLUDGE DISPOSAL[a]

				Total dry solids to disposal		Break down of total dry solids	
Year	Month	Dewatered digested sludge, wet tons per day	Trucking and disposal cost,[b] dollars per day	percent dry solids	tons per day	Digested sludge solids, tons per day	Solids contributed by ferric chloride addition, tons per day[c]
1979	Aug	368	5,520	17.8	65.5		
	Sept	361	5,415	17.9	64.6		
	Oct	345	5,175	16.3	56.2		
	Nov	290	4,350	----	----		
	Dec	289	4,335	16.1	46.5		
1980	Jan	293	4,395	16.0	49.9	44.5	5.4
	Feb	395	5,925	15.8	62.4	56.1	6.3
	Mar	337	5,055	15.6	52.6	46.9	5.7
	Apr	302	4,530	16.7	50.4	45.1	5.3
	May	314	4,710	16.8	52.7	47.0	5.7
	Jun	328	4,920	16.9	55.4	49.9	5.5
	Jul	294	4,410	16.8	49.4		

[a] Data obtained from Mr. Ed Jones, Mr. Steve Bennett, and Blue Plains operating records.

[b] At the time of this report - Sept. 1980 - dewatered digested sludge was being trucked and disposed of on the land for approximately $15/wet ton. This value used for all months evaluated.

[c] It is assumed that all iron (Fe) solids are in the sludge and that 17 percent of the chloride is in the sludge. The remaining 83 percent of the chloride is dissolved in the liquid and leaves in the vacuum filter filtrate.

APPENDIX B

ANALYSIS OF AVERAGE PRIMARY AND SECONDARY SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW

B.1. PRIMARY SLUDGE

TABLE B-1. ANALYSIS OF AVERAGE SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW[a]
(January through June 1980)

Month	Primary sludge production,[b,c] T/MGIF[d]	Primary sludge volatile solids,[b] percent
January	98/331 = 0.296	77
February	116/325 = 0.357	81
March	96/345 = 0.278	78
April	146/332 = 0.440	78
May	119/327 = 0.364	79
June	117/330 = 0.355	85
Average	0.348	79.7
Standard deviation	0.057	2.9

Average dry ton of volatile primary sludge production per MGIF is:

$$(0.348 \pm 0.057)\ \text{T/MGIF}\ (0.797 \pm 0.029)\ \text{percent volatile}$$

$$= (0.348)(0.797) \pm \quad (0.797)^2(0.057)^2 + (0.348)^2(0.029)^2$$

$$= 0.277 \pm 0.047\ \text{VT/MGIF}$$

a All data taken from Appendix A, Table A-1.

b Sludge as withdrawn from the primary clarifiers and pumped to the gravity thickeners.

c It is estimated that grit equal to 1.5 to 2.0 cubic feet per million gallons of influent flow at 100 pounds per cubic foot is included with the primary sludge.

[d] MGIF = million gallons influent flow.
T/MGIF = dry tons of solids per million gallons influent flow.
VT/MGIF = dry tons of volatile solids per million gallons influent flow.

B.2. SECONDARY SLUDGE

B.2.1. Current Conditions

TABLE B-2. ANALYSIS OF AVERAGE SECONDARY SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW[a] (January to June 1980)

Month	Secondary sludge production,[f] T/MGIF	Secondary sludge[b] Volatile solids, percent
January	103/331 = 0.311	66
February	94/325 = 0.289	68
March	82/345 = 0.238	66
April	104/332 = 0.313	66
May	80/327 = 0.245	62
June	101/330 = 0.306	64
Average	0.284	65.3
Standard deviation	0.034	2.1

Average dry tons secondary sludge, volatile solids production per MGIF is:

$$(0.284 \pm 0.034)\ \text{T/MGIF}\ (0.653 \pm 0.021)\ \text{percent volatile}$$

$$= (0.284)(0.653) \pm (0.653)^2(0.034)^2 + (0.284)^2(0.021)^2$$

$$= 0.185 \pm 0.022\ \text{VT/MGIF}$$

[a] All data taken from Appendix A, Table A-1.

[b] Sludge as withdrawn from the secondary clarifiers and pumped to the dissolved air flotation thickeners.

B.2.2. Future Conditions

Increase Phosphorus Removal--
Effluent limitations on phosphorus will be reduced from the current 1.1 mg/l average to the new NPDES permit requirement of 0.53 mg/l. This will be accomplished by increasing the current iron dosage. Past records indicate 3.8 pounds of chemical sludge produced per pound of P removed by iron, therefore, every million gallons of influent will generate an additional:

$$(1.1 - 0.53)\text{mg/l P}\ (8.34)(1\ \text{MG})\ \frac{(3.8\ \text{lbs sludge})}{\text{lb P removed}}\ \frac{(1\ \text{ton})}{2000\ \text{lbs}} = 0.009\ \text{T/MGIF}$$

It is assumed there are no volatile solids.

Multi-media Filters--
Within the year the new multi-media polishing filters will be in operation. It is estimated that these filters will reduce the suspended solids being discharged from an average of 15 to 7.5 mg/l. Therefore, every million gallons of influent will generate an additional:

$$(15.0 - 7.5)\text{mg/l}\ (8.34)(1\ \text{MG})\ \frac{(1\ \text{ton})}{2000\ \text{lbs}} = 0.031\ \text{T/MGIF}$$

It is assumed that 70 percent of the solids will be volatile.

Nitrification--
Within several months the nitrification system will be completely operational. At a minimum, this system will have to reduce current total kjeldahl nitrogen (TKN) levels in the plant effluent from an average of 13.8 mg/l to the NPDES permit level of 5.3 mg/l. Assuming a biological yield coefficient of 0.1 pounds of solids produced per pound of TKN reduced, the additional dry solids generated would be:

$$(13.8 - 5.3)\text{mg/l}\ (8.34)(1\text{MG})\ \frac{(0.1\ \text{lbs solids})}{\text{lb TKN reduced}}\ \frac{(1\ \text{Ton})}{2000\ \text{lbs}} = 0.004\ \text{T/MGIF}$$

It is assumed that 70 percent of the solids will be volatile.

Assuming no change in the standard deviation, the new volatile solids content would be:

$$\frac{(0.185 + 0 + 0.0313(0.7) + 0.0035(0.7)\ (100)}{0.284 + 0.009 + 0.031 + 0.004} = 63.8\ \text{percent}$$

APPENDIX C

ANALYSIS OF GRAVITY THICKENER OPERATION[a]

There are six 65-foot diameter units. Surface area of each unit is 3,318 square feet. Surface area is 19,908 square feet. All operational data taken from Appendix A, Table A-1 and A-2.

TABLE C-1. AVERAGE MONTHLY HYDRAULIC AND SOLIDS LOADING RATES
AVERAGE MONTHLY HYDRAULIC AND SOLIDS LOADING RATES
(January through June 1980)

Month	Hydraulic loading rate, gallons per day per square foot of tank surface	Solids loading rate, gallons per day per square foot of tank surface
January	804	10.7
February	799	12.8
March	789	10.5
April	789	15.5
May	769	12.6
June	804	12.4

TABLE C-2. AVERAGE MONTHLY THICKENING PERFORMANCE
(January through June 1980)

Month	Thickened sludge volume, gallons thickened sludge per million gallons of plant influent	Thickened sludge concentration, percent solids
January	328,000/331 = 991	7.3
February	437,000/325 = 1,345	6.7
March	365,000/345 = 1,058	6.7
April	521,000/332 = 1,569	6.9
May	435,000/327 = 1,330	6.4
June	370,000/330 = 1,121	7.9
Average	1,236	7.0
Standard deviation	218	0.5

No relationship between hydraulic loading rate or solids loading rate versus thickened sludge concentration could be established.

TABLE C-3. ANALYSIS OF EFFECT OF GRAVITY THICKENING ON AVERAGE PRIMARY SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW
(January through June 1980)

Month	Thickened sludge production,[a,b] T/MGIF[c]	Thickened sludge volatile solids,[a] percent
January	105/331 = 0.317	79
February	129/325 = 0.397	75
March	108/345 = 0.313	70
April	158/332 = 0.476	65
May	120/327 = 0.367	66
June	129/330 = 0.391	71
Average	0.377	71
Standard deviation	0.060	5.3

Average VT/MGIF as removed from the gravity thickener.

$$(0.377 \pm 0.060)\ \text{T/MGIF}\ (0.71 \pm 0.053)\ \text{percent volatile}$$

$$= (0.377)(0.71) \pm \quad (0.71)^2(0.060)^2 + (0.377)^2(0.053)^2$$

$$= 0.268 \pm 0.047\ \text{VT/MGIF}$$

[a] Sludge as withdrawn from the gravity thickeners.

[b] It is estimated that grit equal to 1.5 to 2.0 cubic feet per million gallons of influent flow at 100 pounds per cubic foot is included with the primary sludge.

[c] MGIF = million gallons influent flow.
T/MGIF = dry ton of solids per million gallons influent flow.
VT/MGIF = dry ton of volatile solids per million gallons influent flow.

APPENDIX D

ANALYSIS OF DISSOLVED AIR FLOTATION (DAF) THICKENER OPERATION

There are 18, 20-foot wide by 55-foot long (effective length) units. Effective thickening surface area of each unit is 1,100 square feet. Total effective surface area is 19,800 square feet. These units operate continuously at 30 to 34 pounds of secondary solids per day per square foot of effective thickening area. Polymer is required to operate at this loading rate. Polymer usage is 7 to 10 pounds of dry polymer per dry ton of feed solids. All operational data taken from Appendix A, Table A-1 and A-3.

TABLE D-1. AVERAGE MONTHLY THICKENING PERFORMANCE
(January through June 1980)

Month	Thickened sludge volume, gallons thickened sludge per million gallons of plant influent	Thickened sludge concentration, percent solids
January	665,000/331 = 2,009	3.7
February	570,000/325 = 1,754	3.9
March	496,000/345 = 1,438	4.1
April	582,000/332 = 1,753	4.2
May	448,000/327 = 1,370	4.2
June	508,000/330 = 1,539	4.7
Average	1,644	4.1
Standard deviation	239	0.3

Since float concentration is not expected to change in the future, the additional secondary solids generated in the future (Appendix B) will increase thickened sludge volume by approximately the same amount or 15 percent (from 1,644 ± 239 to 1,900 ± 239 gallons per million gallons of influent flow).

At 30 pounds of dry secondary solids per day per square foot of effective thickening area, each DAF can thicken:

$$\frac{\text{30 pounds}}{\text{day-square foot}} \times \frac{\text{1100 square feet}}{\text{unit}} = \frac{\text{33,000 pounds}}{\text{day-unit}}$$

In the future, each million gallons of influent flow will contribute 0.328 ± 0.034 tons (656 ± 68 pounds). Assuming a value of 656 + 68 or 724 pounds per million gallons of influent and an average daily flow of 334 million gallons per day (MGD), then the minimum number of DAF thickeners required to be operating is:

$$\frac{1 \text{ unit}}{33{,}000 \text{ pounds}} \times \frac{724 \text{ pounds}}{1 \text{ MGD}} \times \frac{334 \text{ MGD}}{1} = 8 \text{ units}$$

APPENDIX E

ANALYSIS OF LITHIUM CHLORIDE (LiCl) TRACER STUDIES BLUE PLAINS DIGESTERS

E.1. MECHANICAL DATA

Inside diameter of each digester - 84 feet
Cone depth - 13 feet
Average cylindrical height - 22.1 feet
Theoretical tank volume

$$(42)^2\ (3.1416)(22.1 + 13/3) = 146{,}488 \text{ cubic feet} = 1{,}096{,}000 \text{ gallons}$$

Theoretical weight of digester fluid - 9,140,000 pounds

E.2. TEST PROCEDURE

Each day three digesters were taken off line and 6 pounds (lbs) of LiCl added in approximately four equal fractions at manholes spaced about 25 feet from the center of the digester. On digester No. 1, two of the manholes could not be entered, so one-fourth of the LiCl was added at a manhole located on a line between the two manholes that could not be used, and one-fourth at a location adjacent to the wall of the digester. On digester No. 9, one manhole could not be entered, so one-fourth of the LiCl was added through a manhole at the center of the digester instead. Mixing was accomplished by means of the regular gas-mixing system and with a recycle of approximately 1,000 gallons per minute (GPM) by means of a pump for one hour before the sampling time. Samples were taken before the addition of the LiCl (0 time) and at 3, 6, and 8 hours after the addition. Gas compressors were not available on digesters 9 and 10, so mixing in these digesters was accomplished by means of the 1,000 GPM recycle pump running continuously during the 8-hour test period.

Samples were analyzed for lithium by means of an atomic adsorption instrument, using a graphite furnace. The available volume was calculated from the difference in concentration between the 0-hour and 8-hour samples resulting from the addition of the 6 lb dose of LiCl.

E.3. RESULTS

The results of the tracer studies are presented in Table E-1.

E.4. DISCUSSION

It is apparent that mixing was a problem, as indicated by the fact that in digesters 3, 4, 6, 9, and 11, the 8-hour and 6-hour concentrations differed by more than 10 percent, indicating that equilibrium may not have been reached.

TABLE E-1. LITHIUM TRACER STUDY RESULTS

	Lithium Concentrations, mg/l					Volume,	Usable
						gallons	volume,
Digester	0-hr	3-hr	6-hr	8-hr	8-hr - 0-hr	x 1000	percent
1	0.00	0.95	0.59	0.55	0.55	1,310	119
2	0.16	1.15	1.13	1.09	0.93	775	71
3	0.29	2.11	1.64	1.32	1.69	699	64
4	0.52	4.15	1.82	0.84	0.32	2,251	205
5	0.00	1.50	1.20	1.20	1.20	600	55
6	0.16	1.16	1.14	1.45	1.29	558	51
7	0.39	2.37	1.57	1.43	1.04	693	63
8	0.46	2.17	1.00	1.09	0.63	1,143	104
9	0.03	14.13	3.00	1.37	1.34	538	49
10	0.19	1.09	0.96	0.98	0.79	912	83
11	0.47	18.20	1.50	1.12	0.65	1,108	101
12	0.07	7.15	1.73	1.59	1.52	474	43

The value for digester No. 1 is also suspect. As discussed above, a portion of the LiCl was added near the tank wall of the digester, rather than at the more central addition points used in the other digesters. The very low concentration of LiCl found in digester No. 1 would indicate that mixing was not adequate in the area near the tank wall, and that a portion of the LiCl was not recovered.

The distribution of recycle to the digesters appears to be good, based on the 0-time readings for digesters 4, 8, and 11, the last three tested. The content of LiCl in the digesters before addition of LiCl was 0.52, carried to the digesters by the recycle from the other digesters.

The usable-volume values obtained in these tests exclude portions of the tank volume occupied by grit and also those areas where mixing was so poor that the LiCl never reached them. The results of digester No. 1 indicate that this may be a factor in the areas of the tanks near the walls. It is also of note that 4 of the 9 gas-mixing tubes in the No. 1 digester were inoperative. It should also be noted that a scum layer, which may range from 0 to 12 inches, will also be excluded from the usable volume.

Lithium chloride appears to be quite satisfactory as a tracer material for this test. More reliable results would be obtained if the digesters were kept off-line for a longer period to allow more complete mixing. A test where one slug of LiCl was added to the recycle line would allow a determination of the recycle distribution to all tanks, and a determination of usable volume for the entire system, but a test of this type would also have problems in distinguishing between areas occupied by grit and areas of very poor mixing; to make this determination, longer-term off-line tests are probably required.

The system was simulated by computer analysis, using an average usable digester volume of 700,000 gallons. The values predicted by the simulation agreed very well with the analyzed values through the end of the third day of testing, but the analyzed values were about 20 percent low during the fourth day. It is not known what problems could have caused this discrepancy during the last tests, those of digesters 4, 8, and 11.

E.5. COMPUTER SIMULATION

A computer program was developed to simulate the behavior of lithium chloride during the test period. For the purposes of the simulation, the digesters were divided into four groups of three digesters each, in accordance with the test procedure. Each group was assumed to have a total volume of 2.1 million gallons (700,000 gallons per digester) and a recycle rate of 1,500 gallons per minute, (500 GPM per digester). The flow of fresh sludge into the group and treated sludge out of the tanks was assumed to be 1/15 of the total volume per day. The program calculated the estimated LiCl concentrations each half hour; a subroutine was set up to calculate the changes in concentration that would take place because of the removal of tank contents by the recycle, the addition of recycle sludge that was composed of mixed sludge from all four groups, the removal of treated sludge from the system, and the addition of fresh sludge. The main program in turn applied this subroutine to each group of digesters for each half-hour time period, noting the change in concentration of each group of the combined recycle, taking into account the additions of LiCl to a different group each day, and the fact that one group was off-line for an 8-hour period each day. Results were printed out from 9:00 AM (time of LiCl addition and removal of the group from line) and 5:00 PM (the time that the group was placed back on-line).

The results of the simulation are presented in Table E-2. It can be seen that there was good correlation between the concentrations predicted by the model and the field results until Sunday at 5:00 PM.

Both the 9:00 AM and the 5:00 PM results on Monday, the last day of the test, indicated that actual results were substantially lower than those predicted by the model. We are unable to explain the apparent loss of LiCl from the system indicated by the 5:00 PM, Monday recycle sample. In order to make the computer simulation agree with the field results, it was necessary to postulate a withdrawal of treated sludge and replacement with fresh sludge of 3,300 GPM starting at 5:00 PM on Sunday and continuing through 5:00 PM on Monday or some other addition of several million gallons of fresh sludge to the system. There are no such upsets reported from the plant. One of the digesters did overflow, but this accident took place the next day. The weight of LiCl left (23 lbs) from the original 100 lb drum

TABLE E-2. MODEL SIMULATION OF DIGESTER ANALYSIS

		Lithium concentrations, mg/l				
		Digester Group				
Day	Time	1	2	3	4	Recycle
Friday	5 PM	1.03[a] (1.40)[b]	0	0	0	0
Saturday	9 AM	0.635	0.123 (0.19)	0.123	0.123	0.251
Saturday	5 PM	0.532	1.153 (1.171)	0.171	0.171	0.292-0.51 (0.33)
Sunday	9 AM	0.513	0.821	0.333 (0.34)	0.333	0.500
Sunday	5 PM	0.523	0.740	1.363 (1.37)	0.396	0.553-0.76 (0.75)
Monday	9 AM	0.632	0.740	1.050	0.568 (0.48)	0.747
Monday	5 PM	0.679	0.756	0.974	1.598 (1.02)	0.803 (0.63)

a Concentration derived from model results.

b Concentration derived from field analysis.

checks well with the addition of 6 lbs per digester (72 lbs) and a 2-lb sample of LiCl sent to the lab.

The result of the very low LiCl concentrations found in the system on the last day is that the digesters tested (Nos. 4, 8, and 11) showed high usable volumes of 205, 104, and 101 percent. The good correlation between the computer simulation and actual measured concentrations for the first three days, using an average digester volume of 700,000 gallons, lends credence to the average usable digester volume of 660,000 gallons obtained by the individual measurements if the results for digester 1 (where that part of the LiCl added near the edge of the tank was not recovered) and digesters measured on the last day are ignored.

There is another possible explanation for the divergence of the computer simulation model and the field analyses towards the end of the test period. The 5:00 PM, Monday recycle analysis indicates that the concentration of LiCl is 0.63 mg/l, corresponding to a total weight of 44 lbs of LiCl in the system, as compared to a total weight of 56 lbs predicted by the model. Various explanations are possible:

- o There was a large influx of fresh sludge to the system, flushing out the LiCl. As discussed earlier, this would have had to have been on the order of several million gallons, and no such upset occurred.

- o A mistake was made in the addition of LiCl. As discussed above, a check of the remaining LiCl indicates that the proper amount was added.

- o The average volumes of the digesters are nearer to the theoretical 1,096,000 gallons than the assumed 700,000 gallons. This does not appear reasonable, in that if this assumption is made, all of the analyses for the first three days would have been in poor agreement with the model, and such an assumption does not fit with the individual test results.

- o The missing 12 pounds of LiCl, which cannot be accounted for by the analyses, may be concentrated in areas of poor mixing so that the tracer, or portion of it, never entered the system.

E.6. COMPARISON OF RESULTS WITH CANADIAN TRACER STUDIES

It is of interest to compare results with those obtained by J. Smart (1978) (An Assessment of the Mixing Performance of Several Digesters Using Tracer Response Techniques, Research Publication No. 72, Dec. 1978, Ontario Ministry of the Environment.) He used fluoride as a tracer on 10 anaerobic digesters ranging from 165,900 to 1,690,000 gallons in size, using both mechanical and gas mixing with mixing power ranging from 0.02 to 0.25 HP/1,000 gallons.

Smart found mixing to be a problem, with an average of 55 percent usable volume, as compared to 65 percent found in our study. He found that the tracer concentration leveled out in 4 to 6 hours, which agrees well with our finding that in most cases the tracer leveled out in 4 to 8 hours. He could find no correlation between the amount of dead space and the size, age, condition, applied mixing power, or type of mixing used in the digester.

E.7 CONCLUSIONS

The results are consistent with an average usable volume of about 65 percent of theoretical, with the dead space consisting of the grit layer, the scum layer, and areas where there is little or no mixing.

A tracer study where the tracer is added to the recycle line, thus feeding all 12 digesters at once, will help to resolve some of the questions about short-circuting and total hydraulic detention time.

APPENDIX F

ANALYSIS OF GRIT ENTRY INTO ANAEROBIC DIGESTERS PER MILLION GALLONS OF PLANT INFLUENT FLOW

Based on discussions with anaerobic system operating personnel:

- o It takes 2 to 3 years for a digester to accumulate grit to a point of equilibrium, that is grit no longer accumulates but passes through the digestion tanks.
- o Grit in the tank was estimated to occupy 15 to 20 percent of the total tank volume. Since each tank has a theoretical volume of 146,469 cubic feet, then grit occupies 21,970 to 29,294 cubic feet of tank volume.

Assuming three years for grit build-up within a tank to reach equilibrium then:

$$\frac{\text{21,970 cubic feet of grit}}{\text{(365 day/year) (3 years)}} = \text{20 cubic feet of grit per day}$$

or

$$\frac{\text{29,294 cubic feet of grit}}{\text{(365 days/year) (3 years)}} = \text{26.8 cubic feet of grit per day}$$

Approximately half the daily sludge production goes to anaerobic digestion or the plant influent flow equivalent of 167 million gallons per day (MGD).

There are 12 digestion tanks and it is assumed that sludge is pumped uniformly to all of them, therefore each tank receives raw sluge generated from:

$$\frac{\text{167 MGD}}{\text{12 tanks}} = \text{13.9 MGD of plant influent flow}$$

At 20 cubic feet of grit per day per tank, each MGD of plant influent flow is contributing:

$$\frac{\text{20 cubic feet grit per tank}}{\text{13.9 MGD per tank}} = \text{1.4 cubic feet per MGD}$$

At 100 pounds per cubic foot, 140 pounds or 0.07 tons of dry solids per million gallons of plant influent flow (T/MGIF) is being contributed.

At 26.8 cubic feet it is 2.0 cubic feet per MGD, which at 100 pounds per cubic foot is 200 pounds or 0.1 T/MGIF.

APPENDIX G

ANAEROBIC DIGESTION HEAT AVAILABILITY AND SYSTEM HEAT REQUIREMENTS

G.1. DIGESTER HEATING CAPABILITIES

There are six double inlet, double outlet, 6 inch by 8 inch, tube and tube heat exchangers each with a heated exterior sludge tube surface of 360 square feet. Each unit is specified as originally having a rated output capacity of 3,000,000 BTU per hour. The original design criteria could not be found, but knowing how the heat exchanger supplier would have designed the units, it is believed that the 3,000,000 BTU per hour output was predicated on the following:

- o Sludge at each inlet (there are 12) would be at an average temperature of 80°F at a flow rate of 350 gallons per minute (gpm).
- o Sludge at each outlet (there are 12) would be at an average temperature of 95°F at a flow rate of 350 gpm.
- o Hot water at each inlet (there are 12) would be at 180°F at a flow rate of 200 gpm.
- o Hot water at each outlet (there are 12) would be at 160°F at a flow rate of 200 gpm.

The source of hot water was to be provided by two, 250 horsepower, low pressure steam boilers; one, 150 horsepower, hot water boiler; and engine jacket cooling water.

Through the years since the sludge heating system was first conceived and installed, changes have taken place that have altered the heating capabilities as follows:

- o Sludge at each inlet averages 92 to 95°F at a flow rate of 500 gpm. The higher than expected sludge temperature decreases the expected thermal gradient and therefore decreases heat transfer capabilities. However, the higher flow rate tends to increase heat transfer capabilities.

  The original heat exchanger design would have been predicated on two standard industry assumptions:

  - Sludge flow through a 6-inch sludge tube would be approximately 350 gpm.

- Sludge flow into the heat exchanger would have consisted of 2 parts digested, recirculated sludge at 95°F and 1 part raw sludge at 50°F (assumed winter conditions).

At Blue Plains raw sludge is added after the heat exchangers rather than before. The addition is downstream because of a historical problem with heat exchanger plugging due to rags in the primary sludge. The required screening of plant influent, to eliminate rags in the raw sludge has only recently been accomplished.

Even if all the available raw sludge (approximately 600 gpm) was added upstream, the average inlet temperature would be reduced by 4°F during the winter and by 2°F during the summer. This small decrease is due to the high digested sludge recycle flow, which is 10 times the raw sludge flow instead of the 2 times originally envisioned.

It is not known why the recycle flow rate was increased to 500 gpm per sludge inlet. It has a beneficial effect on the heat exchanger in that the high velocity (5.7 feet per second) through the sludge tubes prevents tube fouling.*

- o Hot water at each inlet is at 195°F. The higher hot water temperature tends to increase the thermal gradient, therefore increasing heat transfer capabilities.

  Plant operating experience indicated that the hot water being utilized could be increased to 195°F without causing any scaling problems within the heat exchanger.*

- o At the present time the source of hot water is provided by two, 250 horsepower, low pressure steam boilers and one, 150 horsepower, hot water boiler. Engine jacket cooling water is no longer available. The 150 horsepower hot water boiler is in poor condition and is scheduled to be replaced within the next couple of years. Assuming an overall efficiency of 80 percent for the steam boilers and heat exchangers, about 13,200,000 BTU per hour are available. The new hot water boiler would add an additional 3,000,000 BTU per hour.

  In addition, the current construction contract calls for linking together the steam boilers in the heat exchanger building with the three 150 horsepower, low pressure steam boilers in the power house and the two 100 horsepower, low pressure steam boilers in the new grit chamber structure being built. These changes would produce an additional 17,000,000 BTU per hour availability, but how much could be used for sludge heating is not known at this time.

G.2. DIGESTION SYSTEM HEAT REQUIREMENTS

Digestion system heat requirements consist of two components: heat required for raw sludge addition and heat required for radiation losses.

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* Inspection of one unit during this study showed no scaling on either sludge or hot water side of 6 inch sludge tube.

G.2.1. <u>Heat Required For Raw Sludge Addition</u>

The amount of heat required for raw sludge addition per million gallons of plant influent flow can be calculated as follows:

$$(\text{gallons of sludge/MGIF})(\text{Density})(T_2 - T_1)$$

where:

gallons of sludge/MGIF = gallons of raw sludge per million gallons of plant influent flow

density = density of the liquid sludge stream, pounds per gallon

T_2 = temperature to which the raw sludge stream is to be raised, °F

T_1 = expected temperature of the raw sludge stream, °F

G.2.2. <u>Heat Required For Primary Sludge Addition Only</u>

From Appendix C, the primary sludge volume expected per MGIF is 1236 ± 218 gallons. The highest value or 1454 gallons is used for the following calculations. Density of primary sludge stream is 8.8 pounds per gallon. Sludge temperature to be raised to 95°F. During the winter season, the coldest primary sludge temperature is 55°F. During summer operation, the primary sludge temperature is 75°F.

Winter operation:

$$\frac{1454 \text{ gallons}}{\text{MGIF}} \times \frac{8.8 \text{ pounds}}{\text{gallon}} \times \frac{(95 - 55)°\text{F}}{1}$$

$$= 511{,}808 \text{ BTU's per MGIF}$$

Summer operation:

$$\frac{1454 \text{ gallons}}{\text{MGIF}} \times \frac{8.8 \text{ pounds}}{\text{gallon}} \times \frac{(95 - 75)°\text{F}}{1}$$

$$= 255{,}904 \text{ BTU's per MGIF}$$

G.2.3. <u>Heat Required For Secondary Sludge Addition Only</u>

From Appendix D, the future secondary sludge volume expected per MGIF is 1900 ± 239 gallons. The highest value or 2139 gallons is used for the following calculations. Density of secondary sludge stream is 8.5 pounds per gallon. Sludge temperature to be raised to 95°F. During the winter season, the coldest secondary sludge temperature expected is 60°F. During summer operation, the secondary sludge temperature expected is 80°F.

Winter operation:

$$\frac{2139 \text{ gallons}}{\text{MGIF}} \times \frac{8.5 \text{ pounds}}{\text{gallon}} \times \frac{(95 - 60)^\circ\text{F}}{1}$$

$$= 636{,}353 \text{ BTU's per MGIF}$$

Summer operation:

$$\frac{2139 \text{ gallons}}{\text{MGIF}} \times \frac{8.5 \text{ pounds}}{\text{gallon}} \times \frac{(95 - 80)^\circ\text{F}}{1}$$

$$= 272{,}723 \text{ BTU's per MGIF}$$

G.2.4. Heat Required For All Sludge Addition

The heat input required for all sludge per MGIF is calculated by totaling the primary and secondary requirements as follows:

	Winter operation, BTU's per MGIF	Summer operation, BTU's per MGIF
Primary sludge	511,808	255,904
Secondary sludge	636,353	272,723
TOTAL	1,148,161	528,627

G.2.5. Heat Required For Conductive/Convective Losses

The amount of heat required for conductive and convective losses from the digestor equipment can be calculated as follows:

$$(U)(\text{Area})(T_3 - T_4)$$

where:

$$U = \frac{1}{\sum 1/C_i + \sum x_i/k_j}$$

C_i = Conductance for a certain thickness of material

$$\frac{\text{BTU}}{\text{hour-square foot-}^\circ\text{F}}$$

x_i = Thickness of material in inches

k_j = Thermal conductivity of material

$$\frac{\text{BTU - inch}}{\text{hour-square foot-}^\circ\text{F}}$$

A = Area of material normal to direction of heat flow in square feet

T_3 = Temperature of inside boundary, °F

T_4 = Temperature of outside boundary, °F

Figure G-1 shows a cross sectional view of one of the 12 existing anaerobic digestion tanks. In the following analysis the assumptions made are:

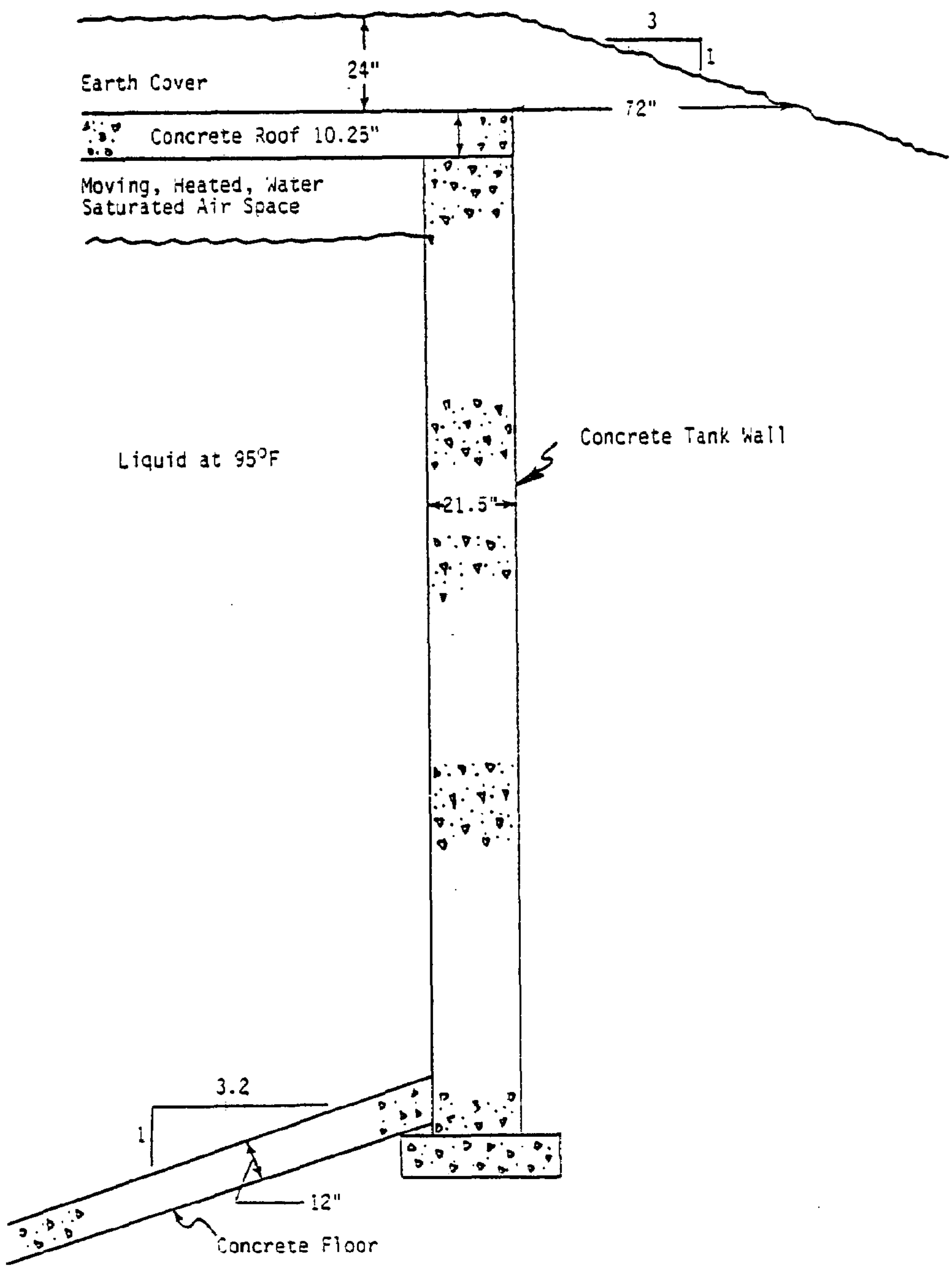

FIGURE G-1. Cross sectional view of existing anaerobic digesters.

1. The liquid contents of the tank are being mixed and maintained at an inside boundry temperature of 95°F.

2. The average (several weeks) coldest outside temperature (winter operation) is 0°F.

3. The average (several weeks) warmest outside temperature (summer operation) is 60°F.

4. All earth is considered to be wet (worst condition).

5. The average temperature of the soil underneath the digestion tanks is 40°F.

G.2.6. Heat Loss Through Roof

T_3 is 95°F; T_4 for winter is 0°F and for summer is 60°F; area normal to heat flow is:

$$\frac{(\pi)(\text{one internal tank diameter})^2\ (12\ \text{tanks})}{4}$$

$$= \frac{(\pi)(84)^2\ (12)}{4} = 66{,}501 \text{ square feet}$$

U can be calculated as follows:

- o liquid-air space interface, $1/C = 0$
- o air space (as indicated), $1/C = 0.5$
- o air space - concrete interface, $1/C = 0.33$
- o concrete 10.25 inches thick, $x/k = \frac{10.25}{12} = 0.854$
- o concrete - earth interface, $1/C = 1.0$
- o wet earth 24 inches thick, $x/k = \frac{24}{16} = 1.5$
- o earth - atmosphere interface, $1/C = 0.167$

$$U = \frac{1}{(0 + 0.50 + 0.33 + 1.00 + 0.167) + (0.854 + 1.50)}$$

$$= \frac{1}{4.35} = \frac{0.23\ \text{BTU}}{\text{hour-square foot-°F}}$$

Winter operation:

$$\frac{(0.23\ \text{BTU})(66{,}501\ \text{square feet})(95-0)\text{°F}}{\text{hour-square foot-°F}} = 1{,}453{,}047 \text{ BTU per hour}$$

Summer operation:

$$\frac{(0.23\ \text{BTU})(66{,}501\ \text{square feet})(95-60)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 535{,}333\ \text{BTU per hour}$$

G.2.7. <u>Heat Loss Through Walls</u>

T_3 is 95°; T_4 for winter is 0°F and for summer is 60°F; area normal to heat flow is:

$$(\pi)(\text{one mean tank diameter})(12\ \text{tanks})(\text{vertical wall height})$$

$$= (\pi)(84 + \frac{21.5}{12})(12)(22) = 71{,}154\ \text{square feet.}$$

All vertical surface is not exposed to the outside. As shown in Figure G-2, the grouping of the existing digestion tanks only allows part of each tank to be exposed to the outside environment, the rest is exposed to each other. The wide dark lines in Figure G-2 indicate the part of the circumference (59 percent) that is considered exposed to the outside. The other 41 percent is exposed to a protected environment with an average year round temperature of 70°F.

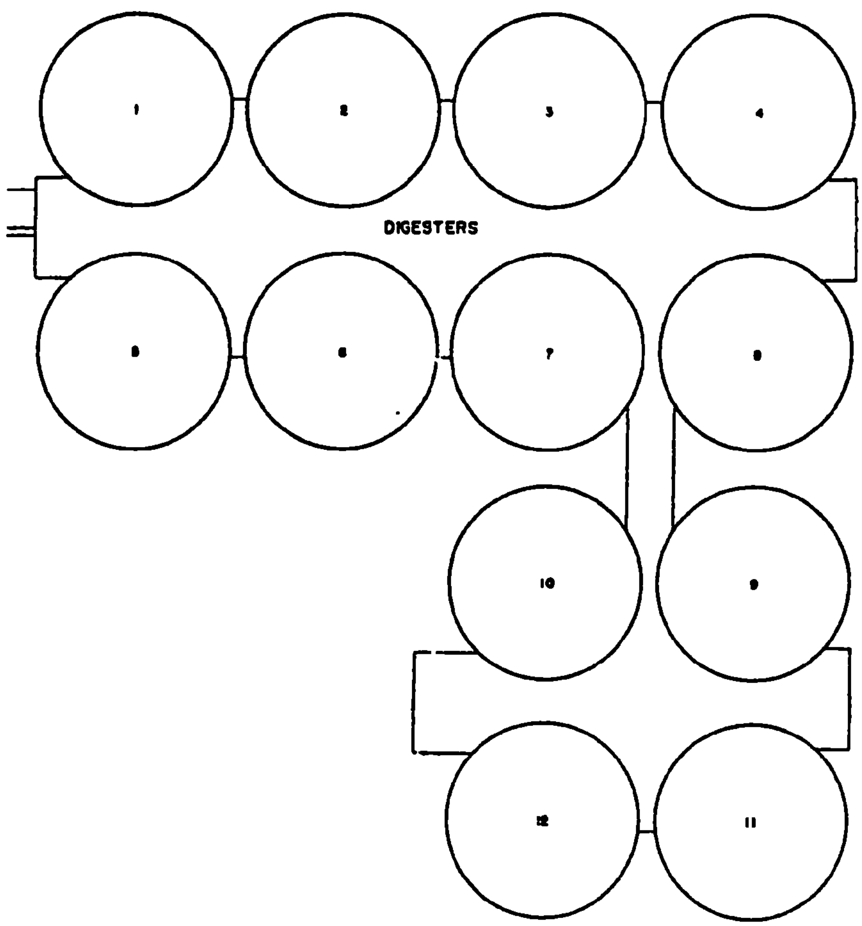

FIGURE G-2. Plan view of existing anaerobic digestion tanks.

U can be calculated as follows:

- o liquid - concrete interface, $1/C = 0$
- o concrete 21.5 inches thick, $x/k = \frac{21.5}{12} = 1.79$
- o concrete - earth interface, $1/C = 1.0$
- o wet earth 72 inches thick, $x/k = \frac{72}{16} = 4.5$*
- o earth - atmosphere interface, $1/C = 0.167$

$$U = \frac{1}{(0 + 1.00 + 0.167) + (1.79 + 4.5)}$$

$$= \frac{1}{7.46} = \frac{0.13 \text{ BTU}}{\text{hour-square feet-}^\circ\text{F}}$$

Winter operation:

$$\frac{(0.13 \text{ BTU})(71{,}154 \text{ square feet})[(0.59)(95-0)^\circ\text{F} + (0.41)(95-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

= 613,276 BTU per hour

Summer operation:

$$\frac{(0.13 \text{ BTU})(71{,}154 \text{ square feet})[(0.59)(95-60)^\circ\text{F} + (0.41)(95-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

= 285,826 BTU per hour

G.2.8. Heat Loss Through Floor

T_3 is 95°F; T_4 is 40°F for both winter and summer; area normal to heat flow is:

$$(\pi)\left(\begin{array}{c}\text{one mean}\\ \text{tank radius}\end{array}\right)\left(\begin{array}{c}\text{one mean}\\ \text{tank radius}\end{array} + \begin{array}{c}\text{vertical height}\\ \text{of cone}\end{array}\right)(12 \text{ tanks})$$

$$= (\pi)\left(42 + \frac{21.5}{(2)(12)}\right)\left(42 + \frac{21.5}{(2)(12)} + 13\right)(12)$$

= 90,391 square feet

---

* In reality, the factor is larger since the earth cover is becoming thicker. In calculating heat loss, the effective protection of an earth cover is never used as more than 10 feet; therefore the maximum value is 7.5.

U can be calculated as follows:

- o liquid - concrete interface, $1/C = 0$
- o concrete 12 inches thick, $x/k = \frac{12}{12} = 1.0$
- o concrete - earth interface, $1/C = 1.0$
- o wet earth 120 inches thick*, $x/k = \frac{120}{16} = 7.5$

$$U = \frac{1}{(0 + 1.0) + (1.0 + 7.5)}$$

$$= \frac{1}{9.5} = \frac{0.11 \text{ BTU}}{\text{hour-square foot-°F}}$$

Winter and summer operation:

$$\frac{(0.11 \text{ BTU})(90{,}391 \text{ square feet})\ (95\text{-}40)\text{°F}}{\text{hour-square foot-°F}} = 546{,}866 \text{ BTU per hour}$$

G.2.9. Heat Losses Through Sludge Piping

All sludge piping involved in transporting sludge from the sludge heat exchangers to the digesters and from the digesters back to the heat exchangers is made of steel pipe of various diameters. It is estimated that approximately 600 linear feet (6,000 square feet of surface area) of this pipe is buried in shallow concrete galleries located between the digestion tanks and heat exchanger complexes. The U factor for this pipe was estimated at 0.32 BTU per hour-square foot-°F. It is estimated that another 3,600 linear feet (10,000 square feet of surface area) is located in several building structures, where the temperature stays about 70°F all year long. The U factor for this pipe is estimated at 1.6 BTU per hour-square foot-°F.

Winter operation:

$$\frac{(0.32 \text{ BTU})(6000 \text{ sq. feet})(95\text{-}0)\text{°F}}{\text{hour-square foot - °F}} + \frac{(1.6 \text{ BTU})(10000 \text{ sq. feet})(95\text{-}70)\text{°F}}{\text{hour-square foot-°F}}$$

$$= 182{,}400 + 400{,}000 = 582{,}400 \text{ BTU per hour}$$

Summer operation:

$$\frac{(0.32 \text{ BTU})(6000 \text{ sq. feet})(95\text{-}60)\text{°F}}{\text{hour-square foot-°F}} + \frac{(1.6 \text{ BTU})(10000 \text{ sq. feet})(95\text{-}70)\text{°F}}{\text{hour-square foot-°F}}$$

$$= 67{,}200 + 400{,}000 = 467{,}200 \text{ BTU per hour}$$

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* In calculating heat loss, the effective protection of an earth cover is never used as more than 10 feet.

G.2.10. Summary of Heat Requirements

	Winter operation	Summer operation
For raw sludge addition in BTU's per MGIF:		
o Primary sludge only	511,808	255,904
o Secondary sludge only	636,353	272,723
TOTAL PRIMARY & SECONDARY SLUDGE	1,148,161	528,627
For system conductive/convective heat loss in BTU's per hour:		
o Roof	1,453,112	535,358
o Walls	613,276	285,826
o Floor	546,866	546,866
o Sludge piping	582,400	467,200
TOTAL HEAT LOSSES	3,195,654	1,835,250

APPENDIX H

ANAEROBIC DIGESTION PIPING SCHEMATIC

In order to perform this study an up-to-date piping schematic of the anaerobic digestion system was required. Enclosed in this appendix are the following three drawings:

Drawing 117-1-1	Existing Piping Schematic Digesters 1-8
Drawing 117-1-2	Existing Piping Schematic Digesters 9-12
Drawing 117-1-3	Existing Exterior Solids Processing Piping

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APPENDIX I

ANALYSIS OF ANAEROBIC SYSTEM VOLATILE MATTER REDUCTION AND GAS PRODUCTION[a]

I.1. VOLATILE MATTER REDUCTION

Table I-1 lists historical volatile matter reduction and several possible related variables. Data are plotted in Figures 5-2, 5-3, and 5-4 in Section 5.3.6 of this report.

TABLE I-1. LISTING OF AVERAGE VOLATILE MATTER REDUCTION AND SEVERAL RELATED VARIABLES

Year	Month	Average volatile matter reduction, percent[b]	Secondary sludge fraction, percent by weight	Volatile matter loading pounds volatile matter per usable cubic foot per day[c]	Hydraulic detention time, days[d]
1979	January	62.9	35	0.15	20.5
	February	64.0	35	0.13	23.9
	March	66.2	19	0.15	22.3
	April	65.0	25	0.16	18.5
	May	65.2	40	0.17	19.3
	June	57.9	35	0.15	19.8
	July	45.3	44	0.12	20.2
	August	52.2	42	0.14	18.8
	September	44.6	52	0.12	18.6
	October	46.9	51	0.12	19.4
	November	--	--	--	26.9
	December	59.6	35	0.10	23.9
1980	January	70.7	23	0.12	23.6
	February	61.6	41	0.14	17.0
	March	60.0	44	0.14	16.4
	April	54.8	46	0.12	18.4
	May	49.7	42	0.12	18.4
	June	57.6	53	0.14	17.6

a All data taken from Appendix A, Table A-4.

b Calculated as follows:

$$\frac{\text{avg tons volatile solids in - avg tons volatile solids out}}{\text{avg tons volatile solids in}} \times 100$$

c Based on the information in Appendix E, it was assumed that only 65 percent of the total 1,757,628 cubic feet of digestion tank volume was usuable.

d Based on the information in Appendix E, it was assumed that only 65 percent of the total 13,152,000 gallons of digestion tank was usuable.

I.2. ANALYSIS OF ANAEROBIC DIGESTION ON AVERAGE VOLATILE MATTER REDUCTION PER MILLION GALLONS OF INFLUENT FLOW*

I.2.1. Current Conditions - All Sludge To Digestion

Primary sludge to anaerobic digestion:

0.377 ± 0.060 T/MGIF**   0.268 ± 0.047 VT/MGIF

Secondary sludge to anaerobic digestion:

0.284 ± 0.034 T/MGIF   0.185 ± 0.023 VT/MGIF

Summation of primary and secondary T/MGIF and VT/MGIF:

$$(0.377 + 0.284) \pm \quad (0.060)^2 + (0.034)^2 \quad = \quad 0.661 \pm 0.069 \quad \text{T/MGIF}$$

$$(0.268 + 0.185) \pm \quad (0.047)^2 + (0.023)^2 \quad = \quad 0.453 \pm 0.052 \quad \text{VT/MGIF}$$

Ratio of secondary sludge to total sludge mass:

$$(0.284) \div (0.661) = 0.43$$

From Figure 5-2, Section 5.3.6, the average volatile solids reduction expected is 56.5 ± 6.5 percent or:

$$(0.453)(0.565) \pm \quad (0.565)^2\,(0.052)^2 + (0.453)^2\,(0.065)^2$$

$$= 0.256 \pm 0.042 \text{ VT/MGIF}$$

The VT/MGIF reduction would be equivalent to the T/MGIF reduced for the total sludge mass.

I.2.2. Future Conditions - All Sludge To Digestion

Primary sludge to anaerobic digestion:

0.377 ± 0.060 T/MGIF   0.268 ± 0.047 VT/MGIF

Secondary sludge to anaerobic digestion:

0.328 ± 0.034 T/MGIF   0.209 ± 0.023 VT/MGIF

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* Data taken from Appendix B, Tables B-1 through B-3.

** T/MGIF = dry ton of solids per million gallons influent flow.
VT/MGIF = dry ton of volatile solids per million gallons influent flow.

Summation of primary and secondary T/MGIF and VT/MGIF:

$$(0.377 + 0.328) \pm \quad (0.060)^2 + (0.034)^2 = 0.705 \pm 0.069 \ \text{T/MGIF}$$

$$(0.268 + 0.209) \pm \quad (0.047)^2 + (0.023)^2 = 0.477 \pm 0.052 \ \text{VT/MGIF}$$

Ratio of secondary sludge to total sludge mass:

$$(0.328) \div (0.705) = 0.465$$

From Figure 5-2, Section 5.3.6, the average volatile solids reduction expected is 53.5 ± 6.5 percent or:

$$(0.477)(0.535) \pm \quad (0.535)^2 (0.052)^2 + (0.477)^2 (0.065)^2$$

$$= 0.255 \pm 0.042 \ \text{VT/MGIF}$$

The VT/MGIF reduction would be equivalent to the T/MGIF reduced for the total sludge mass.

I.3. GAS PRODUCTION

Table I-2 lists historical gas production and several possible related variables. Data are plotted in Figures 5-5, 5-6, and 5-7 in Section 5.3.7 of the report.

TABLE I-2. LISTING OF AVERAGE GAS PRODUCTION AND SEVERAL RELATED VARIABLES

Year	Month	Gas production, cubic feet per pound of volatile matter reduced	Feed solids, percent	Volatile matter to digestion, tons per day	Volatile matter loading, pounds volatile matter per usable cubic feet per day[a]	Secondary sludge fraction, percent by weight	Hydraulic detention time, days[b]
1979	January	8.7	7.1	83.5	0.15	35	20.5
	February	8.8	7.0	74.7	0.13	35	23.9
	March	7.2	7.6	87.2	0.15	19	22.3
	April	8.6	6.3	89.1	0.16	25	18.5
	May	9.9	7.4	94.0	0.17	40	19.3
	June	12.0	7.1	86.4	0.15	35	19.8
	July	14.4	6.0	67.8	0.12	44	20.2
	August	9.2	6.4	77.7	0.14	42	18.8
	September	12.6	5.9	70.6	0.12	52	18.6
	October	13.5	5.6	67.0	0.12	51	19.4
	November	--	--	--	--	--	26.9
	December	9.6	5.4	56.2	0.10	35	23.9
1980	January	11.0	5.9	68.5	0.12	23	23.6
	February	12.8	5.2	78.7	0.14	41	17.0
	March	12.0	5.2	77.8	0.14	44	16.4
	April	12.0	5.3	67.5	0.12	46	18.4
	May	13.8	5.2	66.2	0.12	42	18.4
	June	10.7	5.8	30.6	0.14	53	17.6

[a] Based on the information in Appendix E, it was assumed that only 65 percent of the total 1,757,628 cubic feet of digestion tank volume was usuable.

[b] Based on the information in Appendix E, it was assumed that only 65 percent of the total 13,152,000 gallons of digestion tank volume was usuable.

NOTE: The existing hot water source for the heat exchangers are two steam and one hot water boiler (Appendix G). To allow the boilers to operate at maximum output, approximately 20,250,000 BTU per hour of energy is required. Assuming digester gas at 600 BTU's per cubic foot, then the amount of digester gas needed is:

$$\frac{20,250,000 \text{ BTU per hour required}}{600 \text{ BTU per cubic foot}}$$

$$= 33,750 \text{ cubic feet per hour}$$

$$= 810,000 \text{ cubic feet per day}$$

APPENDIX J

ANALYSIS OF ELUTRIATION SYSTEM OPERATION

Table J-1 lists historical 1980 elutriation underflow to dewatering solids concentration and several possible related variables. Data are plotted in Figures 5-8 through 5-10 in Section 5.4 of this report. All data taken from Appendix A, Table A-5.

TABLE J-1. LISTING OF ELUTRIATION UNDERFLOW TO DEWATERING SOLIDS CONCENTRATION AND SEVERAL POSSIBLE RELATED VARIABLES

Month	Elutriation underflow to dewatering solids concentration, percent solids	Total dry solids to elutriation, tons per day	Volume of digested sludge to elutriation, MGD[a]	Feed solids concentration, percent solids	Solids loading rate, pounds per day per square foot[b]	Ratio of wash water flow to digested sludge flow
January	5.4	45	0.397	2.6	36.7	5.7
February	4.3	53	0.509	2.4	43.2	4.6
March	3.6	53	0.509	2.4	43.3	4.6
April	5.5	54	0.406	2.7	44.1	4.9
May	5.4	58	0.448	3.0	47.4	4.8
June	5.2	70	0.492	3.3	57.1	4.5

a MGD - million gallons per day.

b There are two elutriation tanks each 35-feet wide by 70-feet long operated in series; therefore the first tank receives the entire solids load. Loading rate calculated as follows:

$$\frac{\text{(tons dry solids to elutriation per day) (2000 pounds per ton)}}{\text{(35)(70) square feet of tank}}$$

APPENDIX K

ANALYSIS OF DIGESTED SLUDGE DEWATERING ON AVERAGE SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW*

TABLE K-1.

Month	Ferric chloride usage: Tons per day	Ferric chloride usage: Tons per day of dry solids to filter	Feed solids concentration, percent	Dewatered cake percent solids	Secondary sludge fraction, percent by weight
January	10.2	10.2/54 = 0.189	5.4	16 0	23
February	11 9	11.9/59 = 0.202	4.8	15 8	41
March	10.8	10.8/49 = 0 220	3.6	15.6	44
April	10.0	10 0/46 = 0.217	5.5	16.7	46
May	10.7	10.7/52 = 0.206	5.4	16.8	42
June	10.3	10.3/55 = 0.187	5.2	16 9	53
Average		0.204		16.3	
Standard deviation		0.014		0 6	

The amount of solids contributed by ferric chloride per million gallons of influent flow is estimated below.

The average amount of ferric chloride added per ton of dry solids pumped to the vacuum filters is 0.204 ± 0.014 tons. It is assumed that all iron solids (44 percent of the ferric chloride) and 16.3 ± 0.5 percent of the chloride solids remain in the sludge (the remainder of the chlorides leave with the filtrate). Therefore, the amount of solids contributed by ferric chloride would be:

(0.44 + [0.163 ± 0.006][0.56])(dry weight ferric chloride added)

= (0.53 ± 0.003)(dry weight ferric chloride added)

= (0.53 ± 0.003)(0.204 ± 0.014)(dry solids to dewatering)

= (0.108 ± 0.007)(dry solids to dewatering)

For current conditions, sludge to digestion, each million gallons of influent flow would contribute:

$$(0.661 - 0.256) \pm \sqrt{(0.063)^2 + (0.04)^2}$$

= 0.405 ± 0.075 dry tons to dewatering after anaerobic digestion.

* All data taken from Appendix A, Table A-7 or Appendix I.

The amount of solids contributed by ferric chloride addition would be:

$$(0.108 \pm 0.007)(0.405 \pm 0.075)$$

$$= \quad (0.108)(0.405) \pm \sqrt{(0.405)^2(0.007)^2 + (0.108)^2(0.075)^2}$$

$$= \quad 0.043 \pm 0.009 \text{ T/MGIF*}$$

For future conditions, sludge to digestion, each million gallons of influent flow would contribute:

$$(0.705 - 0.255) \pm \sqrt{(0.063)^2 + (0.04)^2}$$

$=$ 0.450 ± 0.075 dry tons to dewatering after anaerobic digestion.

The amount of solids contributed by ferric chloride addition would be:

$$(0.108 \pm 0.007)(0.450 \pm 0.075)$$

$$= \quad (0.108)(0.450) \pm \sqrt{(0.450)^2(0.007)^2 + (0.108)^2(0.075)^2}$$

$$= \quad 0.049 \pm 0.009 \text{ T/MGIF}$$

For current and future conditions with only primary sludge to digestion, each million gallons of influent flow would contribute:

$$(0.377 - 0.188) \pm \sqrt{(0.055)^2 + (0.035)^2}$$

$=$ 0.189 ± 0.065 dry tons to dewatering after anaerobic digestion.

The amount of solids contributed by ferric chloride addition would be:

$$(0.108 \pm 0.007)(0.189 \pm 0.066)$$

$$= \quad (0.108)(0.189) \pm \sqrt{(0.189)^2(0.007)^2 + (0.108)^2(0.065)^2}$$

$$= \quad 0.020 \pm 0.007 \text{ T/MGIF}$$

For future conditions with only secondary sludge to digestion, each million gallons of influent flow would contribute:

$$(0.328 - 0.073) \pm \sqrt{(0.031)^2 + (0.015)^2}$$

$=$ 0.255 ± 0.034 dry tons to dewatering after anaerobic digestion.

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* T/MGIF = Dry ton of solids per million gallons of influent flow.

The amount of solids contributed by ferric chloride addition would be:

$$(0.108 \pm 0.007)(0.255 \pm 0.034)$$

$$= (0.108)(0.255) \pm \sqrt{(0.255)^2(0.007)^2 + (0.108)^2(0.065)^2}$$

$$= 0.028 \pm 0.010 \text{ T/MGIF}$$

APPENDIX L

ANALYSIS OF RAW SLUDGE DEWATERING ON AVERAGE SLUDGE PRODUCTION PER MILLION GALLONS OF INFLUENT FLOW*

TABLE L-1.

Month	Ferric chloride usage: Tons per day	Ferric chloride usage: Tons per ton of dry solids to filter	Lime usage: Tons per day	Lime usage: Tons per ton of dry solids to filter	Feed solids concentration, percent	Dewatered cake, percent solids	Secondary sludge fraction, percent by weight
January	20.5	20.5/114 = 0.180	55	55/114 = 0.482	4.4	17.2	72
February	17.8	17.8/110 = 0.162	48	48/110 = 0.436	5.2	17.9	44
March	11.5	11.5/75 = 0.153	38	38/75 = 0.507	5.3	21.6	47
April	11.5	11.5/154 = 0.075	48	48/154 = 0.312	5.8	21.1	36
May	11.5	11.5/96 = 0.120	47	47/96 = 0.490	5.5	21.1	38
June	12.5	12.5/108 = 0.116	48	48/108 = 0.444	6.6	22.6	35
Average		0.134		0.445		20.2	
Standard deviation		0.038		0.071		2.2	

The amount of solids currently contributed by lime addition per million gallons of influent flow is estimated below.

The average amount of lime (CaO) added per ton of dry solids pumped to the vacuum filters is 0.445 ± 0.065 tons. It is assumed that all calcium solids (71.5 percent of the CaO) remain in the sludge and that there are 5 percent inerts associated with the high grade CaO utilized. Therefore, the amount of solids contributed by lime addition would be:

$$(0.71 + 0.05)(\text{dry weight CaO added})$$

$$= \quad (0.76)(0.445 \pm 0.071)(\text{dry solids to dewatering})$$

$$= \quad (0.338 \pm 0.054)(\text{dry solids to dewatering})$$

For current conditions, raw sludge to dewatering, the amount of solids contributed by CaO addition per million gallons of influent flow would be:

$$(0.338 \pm 0.054)(0.661 \pm 0.063)$$

$$= \quad (0.338)(0.661) \pm \quad (0.661)^2 (0.054)^2 + (0.338)^2 (0.063)^2$$

$$= \quad 0.223 \pm 0.042 \text{ T/MGIF}^{**}$$

---

* All data taken from Appendix A, Table A-6

** T/MGIF = Dry ton of solids per million gallons of influent flow.

In the future, CaO requirements per ton of dry solids pumped to the vacuum filter are expected to drop by at least 50 percent. This reduction is due to the current addition of lime in excess of conditioning requirements so that a high pH can be developed to meet land trenching requirements. In the near future, trenching will not be utilized and lime requirements will decrease to that required to enhance dewatering operations only. This 50 percent reduction would change the amount of solids contributed by lime addition from [(0.338 ± 0.054)(dry solids to dewatering)] to [(0.169 ± 0.054)(dry solids to dewatering)].

For future conditions, if raw sludge to dewatering, the amount of solids contributed by CaO addition per million gallons of influent flow would be:

$$(0.169 \pm 0.054)(0.705 \pm 0.063)$$

$$= (0.169)(0.705) \pm \sqrt{(0.705)^2 (0.054)^2 + (0.169)^2 (0.063)^2}$$

$$= 0.119 \pm 0.040 \text{ T/MGIF}$$

For future conditions, if only primary sludge to dewatering, the amount of solids contributed by CaO addition per million gallons of influent flow would be:

$$(0.169 \pm 0.054)(0.377 \pm 0.055)$$

$$= (0.169)(0.377) \pm \sqrt{(0.377)^2 (0.054)^2 + (0.169)^2 (0.055)^2}$$

$$= 0.064 \pm 0.022 \text{ T/MGIF}$$

For future conditions, if only secondary sludge to dewatering, the amount of solids contributed by CaO addition per million gallons of influent flow would be:

$$(0.169 \pm 0.054)(0.328 \pm 0.031)$$

$$= (0.169)(0.328) \pm \sqrt{(0.328)^2 (0.054)^2 + (0.169)^2 (0.031)^2}$$

$$= 0.055 \pm 0.018 \text{ T/MGIF}$$

The amount of solids currently contributed by ferric chloride per million gallons of influent flow is estimated below.

The average amount of ferric chloride added per ton of dry solids pumped to the vacuum filters is 0.134 ± 0.038 tons. It is assumed that all iron solids (44 percent of the ferric chloride) and 20.2 ± 2.0 percent of the chloride solids remain in the sludge (the remainder of the chlorides leaves with the filtrate). Therefore, the amount of solids contributed by ferric chloride would be:

$$(0.44 + [0.202 \pm 0.02][0.56](\text{dry weight ferric chloride added})$$

$$= \quad (0.55 \pm 0.011)(\text{dry weight ferric chloride added})$$

$$= \quad (0.55 \pm 0.011)(0.134 \pm 0.038)(\text{dry solids to dewatering})$$

$$= \quad (0.074 \pm 0.021)(\text{dry solids to dewatering})$$

For current conditions, raw sludge to dewatering, the amount of solids contributed by ferric chloride addition per million gallons of influent flow would be:

$$(0.074 \pm 0.021)(0.611 \pm 0.063)$$

$$= \quad (0.074)(0.661) \pm \sqrt{(0.661)^2 (0.021)^2 + (0.074)^2 (0.063)^2}$$

$$= \quad 0.049 \pm 0.015 \text{ T/MGIF}$$

In the future, ferric chloride requirements per ton of dry solids pumped to the vacuum filter are expected to drop at least 15 percent. This reduction is due to the significant reduction in lime requirement brought about by elimination of the trenching operation. This 15 percent reduction would change the amount of solids contributed by ferric chloride addition from [(0.074 ± 0.021)(dry solids to dewatering)] to [(0.063 ± 0.021)(dry solids to dewatering)].

For future conditions, if raw sludge to dewatering, the amount of solids contributed by ferric chloride addition per million gallons of influent flow would be:

$$(0.063 \pm 0.021)(0.705 \pm 0.063)$$

$$= \quad (0.063)(0.705) \pm \sqrt{(0.705)^2 (0.021)^2 + (0.063)^2 (0.063)^2}$$

$$= \quad 0.044 \pm 0.015 \text{ T/MGIF}$$

For future conditions, if only primary sludge to dewatering, the amount of solids contributed by ferric chloride addition per million gallons of influent flow would be:

$$(0.063 \pm 0.021)(0.377 \pm 0.055)$$

$$= \quad (0.063)(0.377) \pm \sqrt{(0.377)^2 (0.021)^2 + (0.063)^2 (0.055)^2}$$

$$= \quad 0.024 \pm 0.009 \text{ T/MGIF}$$

For future conditions, if only secondary sludge to dewatering, the amount of solids contributed by ferric chloride addition per million gallons of influent flow would be:

$$(0.063 \pm 0.021)(0.328 \pm 0.031)$$

$$= (0.063)(0.328) \pm \sqrt{(0.328)^2 (0.021)^2 + (0.063)^2 (0.031)^2}$$

$$= 0.021 \pm 0.007 \text{ T/MGIF}$$

APPENDIX M

SECTION 6.1 SUPPORTING CALCULATIONS

M.1. DIGESTED SLUDGE VACUUM FILTER OPERATION

M.1.1. Current Operation

There are four existing units each having 500 square feet of filtering area. Their yield depends upon the ratio of secondary and primary sludge. In 1974 when the mixture was 40 to 50 percent secondary, the yield averaged 2.41 pounds per square foot per hour.

$$\frac{2.41\text{ pounds}}{\text{square foot-hour}} \times \frac{500\text{ square feet}}{\text{unit}} \times \frac{4\text{ units}}{1} \times \frac{24\text{ hours}}{\text{day}}$$

= 115,680 pounds per day

Assuming an average feed solids concentration of five percent and a density of 8.85 pounds per gallon, the volume of digested sludge capable of being dewatered is:

$$\frac{115{,}680\text{ pounds}}{\text{day}} \times \frac{1\text{ gallon}}{8.85\text{ pounds}} \times \frac{1}{0.05}$$

= 261,423 gallons per day

Assuming an average cake solids of 16.5 percent, the number of wet tons to be hauled away including the ferric chloride ($FeCl_3$) conditioner would be:

$$\left[\frac{115{,}680\text{ pounds sludge}}{\text{day}} + \frac{13{,}180\text{ pounds }FeCl_3}{\text{day}} \right] \times \frac{1\text{ ton}}{2000\text{ pounds}} \times \frac{1}{0.165}$$

= 390.5 wet tons per day

M.1.2. Future Operation

If six new filter units (600 square feet filtering area each with the same yield of 2.41 pounds per square foot per hour) are placed into operation, then:

$$\frac{2.41\text{ pounds}}{\text{square foot-hour}} \times \frac{600\text{ square feet}}{\text{unit}} \times \frac{6\text{ units}}{1} \times \frac{24\text{ hours}}{\text{day}}$$

= 208,224 pounds per day

Assuming same feed solids concentration and density as before, the volume of sludge capable of being dewatered is:

$$\frac{208{,}224 \text{ pounds}}{\text{day}} \times \frac{1 \text{ gallons}}{8.85 \text{ pounds}} \times \frac{1}{0.05}$$

$= 470{,}563$ gallons per day

Assuming an average cake solids of 16.5 percent, the number of wet tons to be hauled away including the ferric chloride ($FeCl_3$) conditioner would be:

$$\left[\frac{208{,}224 \text{ pounds sludge}}{\text{day}} + \frac{23{,}725 \text{ pounds } FeCl_3}{\text{day}}\right] \times \frac{1 \text{ ton}}{2000 \text{ pounds}} \times \frac{1}{0.165}$$

$= 703$ wet tons per day

If the maximum sludge production generated at 334 million gallons per day influent flow was mesophilically digested under the conditions described in Section M.5 of this Appendix, there would be 340,050 pounds per day to dewater. The maximum number of filters required would be:

$$\frac{340{,}050 \text{ pounds}}{\text{day}} \times \frac{1 \text{ filter - day}}{34{,}704 \text{ pounds}}$$

$= 9.8$, or 10 filters required

Assuming an average cake solids at 16.5 percent, the maximum number of wet tons to be hauled away including the ferric chloride ($FeCl_3$) conditioner would be:

$$\frac{340{,}050 \text{ pounds sludge}}{\text{day}} + \frac{38{,}744 \text{ pounds } FeCl_3}{\text{day}} \times \frac{1 \text{ ton}}{2000 \text{ pounds}} \times \frac{1}{0.165}$$

$= 1{,}148$ wet tons per day maximum

M.2. ELUTRIATION OPERATION

In order to generate 115,680 pounds of solids per day in the elutriate underflow at a maximum hydraulic flow rate of 490,000 gallons per day and a density of 8.64 pounds per gallon, the incoming digested solids concentration would have to be:

$$\frac{115{,}680 \text{ pounds}}{\text{day}} \times \frac{1 \text{ gallon}}{8.64 \text{ pounds}} \times \frac{1 \text{ day}}{490{,}000 \text{ gallons}} \times \frac{100}{1}$$

$= 2.7$ percent solids

For any flow less than 490,000 gallons per day, the solids concentration would have to be proportionately higher.

In order to generate 208,224 pounds of solids per day under the same stated conditions, the incoming digested solids concentration would have to be:

$$\frac{208{,}224 \text{ pounds}}{\text{day}} \times \frac{1 \text{ gallon}}{8.64 \text{ pounds}} \times \frac{1 \text{ day}}{490{,}000 \text{ gallons}} \times \frac{100}{1}$$

= 4.9 percent solids

M.3. HEAT EXCHANGER OPERATION*

Available heating capacity limits the amount of sludge that can be hydraulically processed. The maximum flow rate possible based solely on sludge heating capacity is:

1980 winter operation

$$\frac{(9{,}800{,}000 \text{ BTU per hour}) \qquad (24 \text{ hours})}{(8.6 \text{ pounds per gallon})(95 - 55°F)\ (\text{day})}$$

= 683,721 gallons per day

1981 summer operation

$$\frac{(11{,}100{,}000 \text{ BTU per hour}) \qquad (24 \text{ hours})}{(8.6 \text{ pounds per gallon})(95 - 75°F)\ (\text{day})}$$

= 1,548,837 gallons per day

1981 winter operation

$$\frac{(12{,}800{,}000 \text{ BTU per hour}) \qquad (24 \text{ hours})}{(8.6 \text{ pounds per gallon})(95 - 55°F)\ (\text{day})}$$

= 893,023 gallons per day

1982 summer operation

$$\frac{(14{,}100{,}000 \text{ BTU per hour}) \qquad (24 \text{ hours})}{(8.6 \text{ pounds per gallon})(95 - 75°F)\ (\text{day})}$$

= 1,967,442 gallons per day

M.4. DIGESTION TANK OPERATION

The available data indicate that the system can be operated successfully at an average hydraulic residence time of 16 days. The maximum flow rate possible based solely on usable tank capacity is:

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* See Appendix G for discussion of calculation performed.

$$\frac{8{,}545{,}581 \text{ gallons of usable tank volume}}{16 \text{ days}}$$

$= 534{,}099$ gallons per day

If the digestion tanks were cleaned and grit accumulation kept to a minimum, the maximum flow rate possible would increase to:

$$\frac{13{,}147{,}055 \text{ gallons of usable tank volume}}{16 \text{ days}}$$

$= 821{,}691$ gallons per day

If the maximum sludge production generated at 334 million gallons per day influent flow rate was mesophilically digested at a minimum hydraulic detention time of 16 days, and if the existing digesters were clean of grit and grit accumulation was kept to a minimum in the system, the maximum number of digestion tanks required (all the same dimensions) would be:

$$\frac{1{,}200{,}062 \text{ gallons}}{\text{day}} \times \frac{16 \text{ days}}{1} \times \frac{1 \text{ digester}}{1{,}096{,}000 \text{ gallon capacity}}$$

$= 17.5$, or 18 tanks required; six new tanks would be required to meet this condition.

Operating data at Blue Plains indicate that the system can be operated successfully at volatile matter loading ratios averaging 0.16 to 0.17, though gas production seems to deteriorate over 0.14. The maximum volatile matter loading based on 0.16 pounds per usable cubic foot per day would be:

$$\frac{1{,}142{,}458 \text{ cubic feet of usable tank volume}}{1} \times \frac{0.16 \text{ pound volatile matter}}{\text{usable cubic foot}}$$

$= 182{,}793$ pounds volatile matter per day

If the digestion tanks were cleaned and grit accumulation was kept to a minimum, the maximum volatile matter loading based on 0.16 pounds per usable cubic foot per day would be:

$$\frac{1{,}757{,}628 \text{ cubic feet of usable tank volume}}{1} \times \frac{0.16 \text{ pound volatile matter}}{\text{usable cubic foot}}$$

$= 281{,}220$ pounds volatile matter per day

If the maximum sludge production generated at 334 million gallons per day influent flow rate was mesophilically digested at a maximum volatile matter loading of 0.16 pounds per cubic foot per day, and if the existing digesters were clean of grit and grit accumulation was kept to a minimum in the system, the maximum number of digestion tanks required (all the same dimensions) would be:

$$\frac{362{,}022 \text{ pounds volatile matter}}{\text{day}} \times \frac{1 \text{ cubic foot - day}}{0.16 \text{ pounds volatile matter}} \times \frac{1 \text{ tank}}{146{,}469 \text{ cubic feet}}$$

$= 15.4$, or 16 tanks required; hydraulic flow rate governs.

M.5. GALLONS OF PRIMARY AND SECONDARY SLUDGE PUMPED TO DIGESTION

If the existing digestion system was processing all the Blue Plains sludge, the ratio of thickened secondary sludge volume to thickened primary sludge volume would be 1.54:1. Using this ratio for allocating the 490,00 gallons gives:

	Gallons per day	Pounds per day	Volatile Pounds per day
Primary sludge	192,913	118,834	84,373
Secondary sludge	297,087	103,683	66,254
TOTAL	490,000	222,517	150,627

The following constraints apply to the above calculations:

- o Total flow to digestion system to be no greater than 490,000 gallons per day - ok
- o Minimum ratio of secondary sludge mass to total sludge mass to be 0.35

  $$\frac{\text{103,683 pounds secondary sludge per day}}{\text{222,517 pounds total sludge per day}} = 0.47 \text{ - ok}$$

- o Concentration of feed solids into digestion tanks not to exceed six percent

  $$\frac{\text{(222,517 pounds total sludge per day)(100)}}{\text{(8.6 pounds per gallon)(490,000 gallons per day)}} = 5.3\% \text{ - ok}$$

- o Concentration of solids into elutriation system from digestion tanks to exceed 2.7 percent

  From Figure 5-2, at a secondary:total mass ratio of 0.46 an average volatile matter reduction of 53.5 percent is expected.

$$\frac{\text{[(222.517 pounds of sludge per day) - 0.535(150.627 pounds volatile sludge per day)](100)}}{\text{(8.50 pounds per gallon)(490,000 gallons per day)}}$$

$= 3.4$ percent - ok

- o Sludge mass from digestion tank not to exceed 115,680 pounds per day using the existing drum filters for dewatering and 208,234 pounds per day using the six new belt type filters for dewatering.

222,517 pounds of sludge per day - (0.535)(150,627 pounds volatile sludge per day)

= 141,932 pounds per day - not ok for current operation of drum filters; ok for future operation with new vacuum filters.

M.6. GAS PRODUCTION

M.6.1. 490,000 Gallon Per Day Sludge Processing Rate

Under the constraints imposed in developing the 490,000 gallon per day flow rate, gas production is expected to range from 12 to 14 cubic feet per pound of volatile matter reduced.

Volatile matter reduced:

$$(0.535)(150{,}627) \;=\; 80{,}585 \text{ pounds per day}$$

$$\frac{80{,}585 \text{ pounds}}{\text{day}} \times \frac{12 \text{ cubic feet}}{\text{pound}} = 967{,}020 \text{ cubic feet per day}$$

$$\frac{80{,}585 \text{ pounds}}{\text{day}} \times \frac{14 \text{ cubic feet}}{\text{pound}} = 1{,}128{,}190 \text{ cubic feet per day}$$

Gas consumption to meet winter and summer sludge heating requirements are:

Winter operation:

$$\frac{\dfrac{490{,}000 \text{ gallons}}{\text{day}} \times \dfrac{8.6 \text{ pounds}}{\text{gallon}} \times \dfrac{(95-55)^{\circ}\text{F}}{1} + \dfrac{24 \text{ hours}}{\text{day}} \times \dfrac{3{,}195{,}654 \text{ BTU radiation loss}}{\text{hour}}}{\dfrac{600 \text{ BTU}}{\text{cubic foot of gas}} \times \dfrac{0.8 \text{ heat efficiency factor}}{1}}$$

$$= \;510{,}949 \text{ cubic feet per day}$$

Summer operation:

$$\frac{\dfrac{490{,}000 \text{ gallons}}{\text{day}} \times \dfrac{8.6 \text{ pounds}}{\text{gallon}} \times \dfrac{(95-75)^{\circ}\text{F}}{1} + \dfrac{24 \text{ hours}}{\text{day}} \times \dfrac{1{,}835{,}250 \text{ BTU radiation loss}}{\text{hour}}}{\dfrac{600 \text{ BTU}}{\text{cubic foot of gas}} \times \dfrac{0.8 \text{ heat efficiency factor}}{1}}$$

$$= \;267{,}346 \text{ cubic feet per day}$$

Average expected excess:

$$\frac{(967{,}020 + 1{,}128{,}190) \text{ cubic feet}}{2 \text{ day}} - \frac{(510{,}949 + 267{,}346) \text{ cubic feet}}{2 \text{ day}}$$

$$= \;658{,}458 \text{ cubic feet per day available to be sold}$$

M.6.2. Maximum Gallon Per Day Sludge Processing Rate

Again, gas production is expected to range from 12 to 14 cubic feet per pound of volatile matter reduced.

Volatile matter reduced:

$$(0.535)(319{,}030) = 170{,}681 \text{ pounds per day}$$

$$\frac{170{,}681 \text{ pounds}}{\text{day}} \times \frac{12 \text{ cubic feet}}{\text{pound}} = 2{,}048{,}172 \text{ cubic feet per day}$$

$$\frac{170{,}681 \text{ pounds}}{\text{day}} \times \frac{14 \text{ cubic feet}}{\text{pound}} = 2{,}389{,}534 \text{ cubic feet per day}$$

Gas consumption to meet winter and summer sludge heating requirements (based on 18 digestion tanks) would be:

Winter operation:

$$\frac{\dfrac{1{,}200{,}062 \text{ gallons}}{\text{day}} \times \dfrac{8.6 \text{ pounds}}{\text{gallon}} \times \dfrac{(95-55)^{\circ}\text{F}}{1} + \dfrac{24 \text{ hours}}{\text{day}} \times \dfrac{5{,}272{,}829 \text{ BTU radiation loss}}{\text{hour}}}{\dfrac{600 \text{ BTU}}{\text{cubic feet of gas}} \times \dfrac{0.8 \text{ heat efficiency factor}}{1}}$$

$$= 1{,}123{,}686 \text{ cubic feet per day}$$

Summer operation:

$$\frac{\dfrac{1{,}200{,}062 \text{ gallons}}{\text{day}} \times \dfrac{8.6 \text{ pounds}}{\text{gallon}} \times \dfrac{(95-75)^{\circ}\text{F}}{1} + \dfrac{24 \text{ hours}}{\text{day}} \times \dfrac{3{,}028{,}162 \text{ BTU radiation loss}}{\text{hour}}}{\dfrac{600 \text{ BTU}}{\text{cubic feet of gas}} \times \dfrac{0.8 \text{ heat efficiency factor}}{1}}$$

$$= 581{,}430 \text{ cubic feet per day}$$

Average expected excess:

$$\frac{(2{,}048{,}172 + 2{,}389{,}534)}{2}\frac{\text{cubic feet}}{\text{day}} - \frac{(1{,}123{,}686 + 581{,}430)}{2}\frac{\text{cubic feet}}{\text{day}}$$

$$= 1{,}366{,}295 \text{ cubic feet per day available to be sold}$$

APPENDIX N

SECTION 6.2 SUPPORTING CALCULATIONS

N.1. DIGESTED SLUDGE VACUUM FILTER OPERATION

Each vacuum filter has 600 square feet of filtering area. Assuming a minimum yield for this sludge, thermophilically digested, of 3.0 pounds per hour per square foot, each filter would be able to dewater:

$$\frac{3.0 \text{ pounds}}{\text{square foot-hour}} \times \frac{600 \text{ square feet}}{\text{filter}} \times \frac{24 \text{ hours}}{\text{day}}$$

$= 43{,}200$ pounds per day

As indicated elsewhere in this Appendix, if the maximum sludge production generated at 334 million gallons per day influent flow rate was thermophilically digested, there would be 340,050 pounds per day to dewater. The maximum number of filters required would be:

$$\frac{340{,}050 \text{ pounds}}{\text{day}} \times \frac{1 \text{ filter-day}}{43{,}200 \text{ pounds}}$$

$= 7.9$, or 8 filters required

Assuming an average cake solids of 16.5 percent, the maximum number of wet tons to be hauled away including the polymer conditioner would be:

$$\frac{340{,}050 \text{ pounds sludge}}{\text{day}} + \frac{680 \text{ pounds polymer solids}}{\text{day}} \times \frac{1 \text{ ton}}{2000 \text{ pounds}} \times \frac{1}{0.165}$$

$= 1{,}033$ wet tons per day maximum

N.2. ELUTRIATION (COOLING) OPERATION

Sludge elutriation is not required, instead, the elutriation tanks would be utilized as sludge cooling tanks. The system would still be operated as a two-stage series, co-current, washing operation but under different washing rates. The maximum hydraulic flow through digestion at 334 million gallon per day influent flow rate is 1,200,062 gallons per day. The temperature of digested sludge is 122°F. Cooling water on a 1:1 flow basis would be mixed with digested sludge before entering the elutriation tanks. The warmest cooling water is assumed to be 60°F. In the first stage, 1,200,062 gallons of sludge at 122°F is mixed with 600,031 gallons of wash water at 60°F. Under these conditions, the minimum sludge discharge temperature would be:

$$\frac{(1{,}200{,}062 \text{ gallons})(122°\text{F}) + (600{,}031 \text{ gallons})(60°\text{F})}{(1{,}200{,}062 + 600{,}031) \text{ gallons}} = 101.3°\text{F}$$

If we assume that the sludge concentrates to 75 percent of its original volume, is pumped to the second stage, and blended with 600,031 gallons of wash water at 60°F, the minimum sludge temperature discharge to dewatering would be:

$$\frac{(1{,}200{,}062 \text{ gallons})(0.75)(101.3°F) + (600{,}031 \text{ gallons})(60°F)}{[(1{,}200{,}062)(0.75) + 600{,}031] \text{ gallons}} = 84.8°F$$

NOTE: It may be economically attractive to build a tube and tube heat exchanger for raw sludge going to digestion and digested sludge leaving digestion.

N.3. EFFECT ON SYSTEM HEATING CAPABILITIES DUE TO INCREASE IN SLUDGE OPERATING TEMPERATURE*

N.3.1. Heat Required for Primary Sludge Addition Only

Winter operation:

$$\frac{1454 \text{ gallons}}{\text{MGIF**}} \times \frac{8.8 \text{ pounds}}{\text{gallon}} \times \frac{(122 - 55)°F}{1}$$

$= 857{,}278$ BTU's per MGIF

Summer operation:

$$\frac{1454 \text{ gallons}}{\text{MGIF}} \times \frac{8.8 \text{ pounds}}{\text{gallon}} \times \frac{(122 - 75)°F}{1}$$

$= 601{,}374$ BTU's per MGIF

N.3.2. Heat Required for Secondary Sludge Addition

Winter operation:

$$\frac{2139 \text{ gallons}}{\text{MGIF}} \times \frac{8.5 \text{ pounds}}{\text{gallon}} \times \frac{(122 - 60)°F}{1}$$

$= 1{,}127{,}253$ BTU's per MGIF

Summer operation:

$$\frac{2139 \text{ gallons}}{\text{MGIF}} \times \frac{8.5 \text{ pounds}}{\text{gallon}} \times \frac{(122 - 80)°F}{1}$$

$= 763{,}623$ BTU's per MGIF

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* Assumptions and calculations are as per Appendix G except digestion operating temperature at 122°F rather than 95°F.

** MGIF - million gallons influent flow.

N.3.3. Heat Required for Conductive/Convective Losses

Roof - winter operation:

$$\frac{(0.23\ \text{BTU})(66{,}501\ \text{square feet})(122 - 0)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 1{,}866{,}018\ \text{BTU per hour}$$

Roof - summer operation:

$$\frac{(0.23\ \text{BTU})(66{,}501\ \text{square feet})(122 - 60)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 948{,}304\ \text{BTU per hour}$$

Walls - winter operation:

$$\frac{(0.13\ \text{BTU})(71{,}154\ \text{square feet})[(0.59)(122-0)^\circ\text{F} + (0.41)(122-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

$$= 863{,}027\ \text{BTU per hour}$$

Walls - summer operation:

$$\frac{(0.13\ \text{BTU})(71{,}154\ \text{square feet})[(0.59)(122-60)^\circ\text{F} + (0.41)(122-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

$$= 535{,}576\ \text{BTU per hour}$$

Floor - summer and winter operation:

$$\frac{(0.11\ \text{BTU})(90{,}391\ \text{square feet})(122-40)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 815{,}327\ \text{BTU per hour}$$

Sludge piping - winter operation:

$$\frac{(0.32\ \text{BTU})(6000\ \text{square ft})(122-0)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} + \frac{(1.6\ \text{BTU})(10{,}000\ \text{square ft})(122-70)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}}$$

$$= 234{,}240 + 832{,}000 = 1{,}066{,}240\ \text{BTU per hour}$$

Sludge piping - summer operation:

$$\frac{(0.32\ \text{BTU})(6000\ \text{square ft})(122-60)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} + \frac{(1.6\ \text{BTU})(10000\ \text{square ft})(122-70)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}}$$

$$= 119{,}040 + 832{,}000 = 951{,}040\ \text{BTU per hour}$$

N.3.4. Summary of Heat Requirements

	Winter operation	Summer operation
For raw sludge addition in BTU's per MGIF		
Primary sludge only	857,278	601,374
Secondary sludge only	1,127,253	763,623
TOTAL PRIMARY & SECONDARY SLUDGE	1,984,531	1,364,997

For system heat loss in BTU's per hour

	Winter operation	Summer operation
° Roof	1,866,018	948,304
° Walls	863,027	535,576
° Floor	815,327	815,327
° Sludge piping	1,066,240	951,040
TOTAL HEAT LOSSES	4,610,612	3,250,247

Based on a 334 million gallon per day (MGD) of influent flow, the total digestion sludge heating requirements would be:

Winter operation:

$$\frac{(1{,}984{,}531\ \text{BTU's})(334\ \text{MGIF})}{(\text{MGIF})\qquad(\text{day})} + \frac{(4{,}610{,}612\ \text{BTU})(24\ \text{hours})}{(\text{MGIF})\qquad(\text{day})}$$

$= 7.8 \times 10^8$ BTU's per day (32.2×10^6 BTU per hour)

Summer operation:

$$\frac{(1{,}364{,}997\ \text{BTU's})(334\ \text{MGIF})}{(\text{MGIF})\qquad(\text{day})} + \frac{(3{,}250{,}247\ \text{BTU})(24\ \text{hours})}{(\text{hour})\qquad(\text{day})}$$

$= 5.3 \times 10^8$ BTU's per day (22.2×10^6 BTU per hour)

N.4. EFFECT ON SLUDGE PIPING DUE TO INCREASE IN SLUDGE OPERATING TEMPERATURE

All sludge piping involved in transporting sludge from the sludge heat exchangers to the digesters and from the digesters back to the heat exchangers is made of steel, located in pipe galleries protected from outside weather conditions, and not ridgidly anchored at the ends. Any increase in sludge operating temperature will cause a corresponding increase in pipe expansion. Linear expansion would be of primary concern at this installation.

The additional linear expansion of the sludge piping due to increasing the sludge temperature from the existing operating temperature to a higher operating temperature can be calculated using the equation.

$$L = (\alpha)(L)(\Delta T)$$

where

L = the change in length of the piping system being considered, inches.

α = the coefficient of linear expansion, for steel it is 78×10^{-6} inches per foot per °F

L = length of piping being considered, feet

Δ T = increase in temperature, °F

The sludge piping involved currently operates at a temperature of 90°F to 93°F. Operating at a maximum sludge temperature of 130°F, the maximum increased temperature, T, would be 40°F.

Maximum linear expansion would occur on the longest pipe segment. The longest pipe segment is 400 feet long and is located on the north side of the digestion facility where it carries heated sludge to the anaerobic digesters.

The maximum increased linear expansion, L, is calculated to be

$$\frac{78 \times 10^{-6} \text{ inches}}{\text{foot - °F}} \times 400 \text{ feet} \times 40\text{°F}$$

or 1.25 inches

Based on calculations for expansion of all other pipe segments and best engineering judgement, the increase in temperature and the resultant pipe expansion will not adversely effect the sludge piping involved.

N.5. EFFECT ON ANAEROBIC DIGESTION TANKS DUE TO INCREASE IN SLUDGE OPERATING TEMPERATURE

Increasing the sludge temperature within the existing concrete digestion tanks will create increased stresses on the outer face of the digestion tank walls. The amount of stress developed in the concrete and the amount of reinforcing steel required needs to be calculated and compared to the existing situation to make sure a structural failure would not occur.

Figures N-1 and N-2 show the situation that would occur on the existing structure assuming that the maximum sludge temperature is 130°F and the coldest ambient air temperature is 0°F.

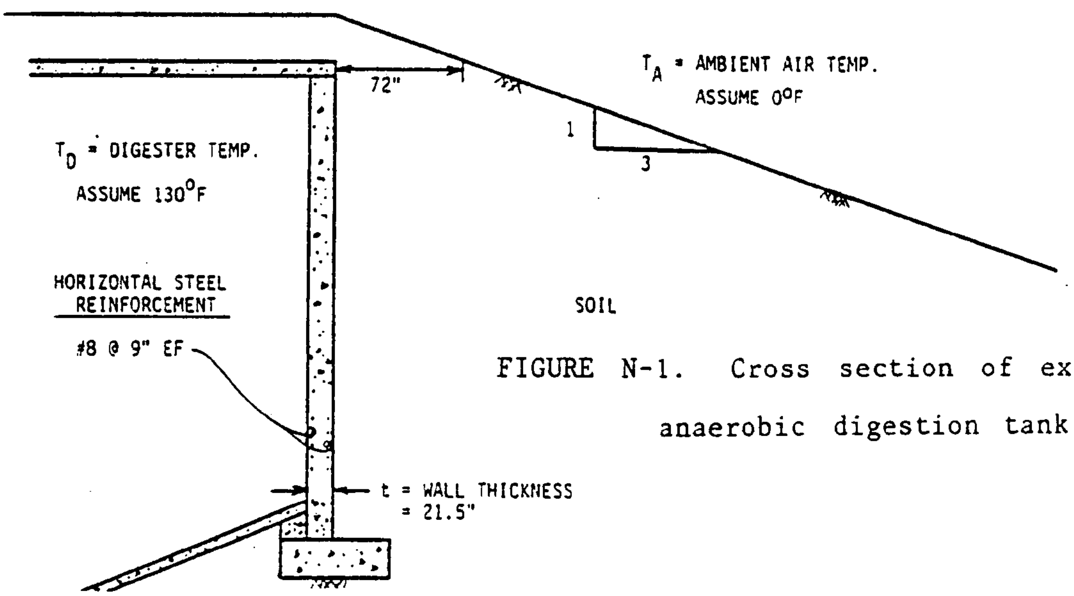

FIGURE N-1. Cross section of existing anaerobic digestion tank wall.

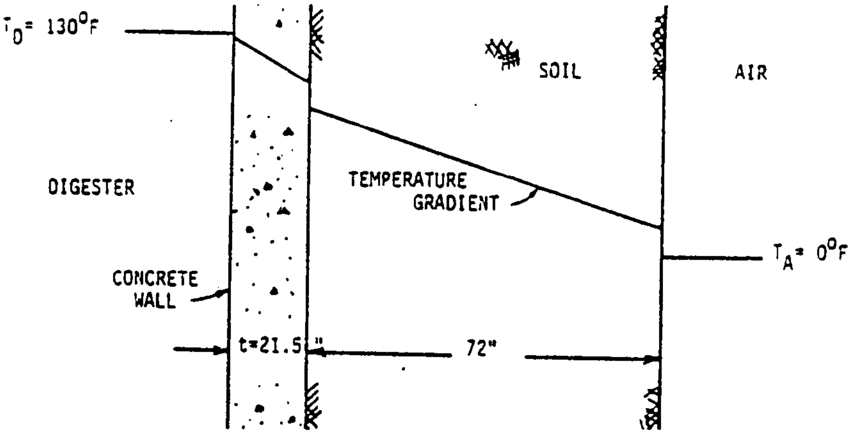

FIGURE N2. Assumed temperature gradient through digester wall and soil.

As indicated in Figures N-1 and N-2, the existing tank walls are 21.5 inches thick with #8 rebar placed every 9 inches. In addition, the entire structure is buried; the minimum soil cover on the side walls being 72 inches.

The stress in the outer concrete wall is calculated on the following two pages.

Temperature difference through the wall, T, can be found by the following equations.

$$T = (T_D - T_A)\ \frac{t/K}{1/K}$$

and

$$1/K = 1/S + t/K + t_1/K_1$$

where:

T_D = digester temperature (°F)

T_A = ambient air temp (°F)

K = coefficient of conductivity for concrete

K_1 = coefficient of conductivity for soil

t = thickness of concrete (inches)

t_1 = thickness of soil (inches)

S = outside surface coefficient

$$T = (130-0)\ \frac{\frac{21.5}{12}}{\frac{1}{6} + \frac{21.5}{12} + \frac{72}{8}} = 22°F$$

Section 18. Temperature Stresses in Cylindrical Tank Walls [a]

Tanks containing hot liquids are subject to temperature stresses. Assume that the temperature is T_1 in the inner face, T_2 in the outer face, and that the temperature decreases uniformly from inner to outer face, $T_1 - T_2$ being denoted as T. Fig 37 shows a segment of a tank wall in two positions, one before and one after a uniform increase in temperature. The original length of the arc of the wall has been increased, but an increase that is uniform throughout will not create any stresses as long as the ring is supposed to be free and unrestrained at its edges. It is the temperature differential only, T, which creates stresses.

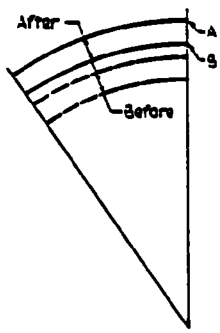

FIG. 37

The inner fibers being hotter tend to expand more than the outer fibers, so if the segment is cut loose from the adjacent portions of the wall, Point A in Fig 38 will move to A', B will move to B', and section

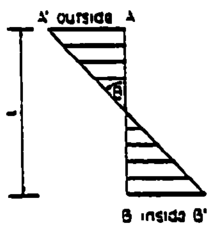

FIG. 38

AB, which represents the stressless condition due to a uniform temperature change throughout, will move to a new position $A'B'$. Actually the movements from A to A' and B to B' are prevented since the circle must remain a circle, and stresses will be created that are proportional to the horizontal distances between AB and $A'B'$.

It is clear that $AA' = BB' =$ movement due to a temperature change of $\frac{1}{2}T$ or when ϵ is the coefficient of expansion, that

$$AA' = BB' = \tfrac{1}{2}T \times \epsilon \text{ per unit length of arc,}$$

and

$$\theta = \frac{AA'}{\frac{1}{2}t} = \frac{T \times \epsilon}{t}$$

In a homogeneous section, the moment M required to produce an angle change θ in an element of unit length may be written as

$$M = EI\theta$$

Eliminating θ gives

$$M = \frac{EI \times T \times \epsilon}{t}$$

The stresses in the extreme fibers created by M are

$$f = \frac{M}{I} \times \frac{t}{2} = \tfrac{1}{2}E \times T \times \epsilon$$

The stress distribution across the cross section is as indicated in Fig. 38. The stresses are numerically equal at the two faces but have opposite signs. Note that the equation applies to uncracked sections only, and that this procedure of stress calculation is to be considered merely as a method by which the problem can be approached. The variables E and I in the equations are uncertain quantities. E may vary from 1,500,000 up to 4,500,000 p.s.i., and I may also vary considerably because of deviations from the assumption of linear relation between stress and strain. Finally, if the concrete cracks, M can no longer be set equal to $EI\theta$, nor f equal to $\frac{M}{I} \times \frac{t}{2}$. As a result, the equation $f = \frac{1}{2}ET\epsilon$ is to be regarded as merely indicative rather than formally correct.

The value of ϵ may be taken as 0.000006, and for the purpose of this problem choose $E = 1{,}500{,}000$ p.s.i. Then $E \times \epsilon = 9$ and $f = 4.5T$

The value of T is the difference between temperatures in the two surfaces of the concrete which may be computed from the temperature of the stored liquid and the outside air.

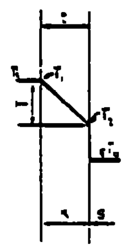

FIG. 39

[a] Circular Concrete Tanks Without Prestressing. Portland Cement Association. 5420 Old Orchard Road, Skokie, Ill. 60076.

When the flow of heat is uniform from the inside to the outside of the wall section in Fig. 39, the temperature difference, $T = T_1 - T_2$, is smaller than the difference, $T_i - T_o$, between the inside liquid and the outside air. Standard textbooks give

$$T = (T_i - T_o)\frac{t/k}{1/k}$$

in which

$$\frac{1}{k} = \frac{1}{s} + \frac{t}{k} + \frac{t_1}{k_1}$$

- k = coefficient of conductivity of stone or gravel concrete = 12 B.t.u. per hour per sq.ft. per deg. F. per in. of thickness
- k_1 = coefficient of conductivity of insulating material
- t = thickness of concrete in in.
- t_1 = thickness of insulating material
- s = outside surface coefficient = 6 B.t.u. per hour per sq.ft. per deg. F

Assuming an uninsulated wall

$$T = (T_i - T_o)\frac{\frac{t}{12}}{\frac{1}{6} + \frac{t}{12}} = (T_i - T_o)\frac{t}{2 + t}$$

Consider a tank with wall thickness $t = 10$ in. which holds a liquid with a temperature $T_i = 120$ deg F while the temperature of the outside air $T_o = 30$ deg. F Then

$$T = (120 - 30) \times \frac{10}{2 + 10} = 75 \text{ deg. F}$$

and

$$f = 4.5T = 4.5 \times 75 = 375 \text{ p.s.i.}$$

The stress of $f = 375$ p.s.i. is tension in the outside and compression in the inside face. If the uniformly distributed ring tension due to load in the tank is, say, 300 p.s.i., the combined stress will be:

Outside fiber: 300 + 375 = 675 p.s.i. (tension)
Inside fiber: 300 − 375 = −75 p.s.i. (compression)

In reality, too much significance should not be attached to the temperature stress computed from the equation derived. The stress equation is developed from the strain equation, $AA' = \frac{1}{2} Te$, based on the assumption that stress is proportional to strain. This assumption is rather inaccurate for the case under discussion. The inaccuracy may be rectified to some extent by using a relatively low value for E_c, such as $E_c = 1{,}500{,}000$ p.s.i., which is used in this section. An even lower value may be justified.

As computed in the example, a temperature differential of 75 deg F. gives a stress of 375 p.s.i. in the extreme fiber. This is probably more than the concrete can take in addition to the regular ring tension stress without cracking on the colder surface. The temperature stress may be reduced by means of insulation, which serves to decrease the temperature differential, or additional horizontal reinforcement may be provided close to the colder surface. A procedure will be illustrated for determination of temperature steel. It is not based upon a rigorous mathematical analysis but will be helpful as a guide and as an aid to engineering judgment.

It is proposed to base the design on the moment derived in this section, $M = E_c I T e/t$, in which the value of E_c is taken as 1,500,000 p.s.i. If I is taken for a section 1 ft. high, I equals t^3, and M is the moment per ft. Then

$M = 1{,}500{,}000 \times t^2 \times T \times 0.000006 = 9t^2T$ in.lb. per ft. in which

t = thickness of wall in in.
T = temperature differential in deg. F.

The area of horizontal steel at the colder face computed as for a cracked section is

$$A_s = \frac{M}{\frac{7}{8} f_s d} = \frac{9t^2T}{17{,}500d} \text{ sq.in. per ft.}$$

For example, assume $t = 15$ in., $T = 75$ deg. F , and $d = 13$ in., which gives

$$A_s = \frac{9 \times 15^2 \times 75}{17{,}500 \times 13} = 0.67 \text{ sq.in. per ft.}$$

This area is in addition to the regular ring steel.

The stress, f, on the concrete can be determined by the equation:

$$f = 4.5\ T = 4.5\ (22°F) = 99\ \text{psi}$$

The maximum stress on the wall should be less than 300 psi as dictated by common engineering practice. Since the temperature stress is only 99 psi, the concrete will not crack.

The amount of reinforcing steel required to resist the temperature stress can be calculated from the following equation:

$$A_S = \frac{9t^2T}{17{,}500d}$$

where A_S = cross sectional area of steel per foot of wall - square inches (sq. in.)

d = effective concrete thickness (21.5" - 2(3") = 15.5")

$$A_S = \frac{9(21.5)^2(22)}{17{,}500\ (15.5)} = 0.34\ \text{sq. in.}$$

The maximum temperature stress occurs in the outer wall face. The amount of steel in that face is one #8 bar every 9 inches. Since a #8 bar equals 1 inch diameter or 0.79 sq. in., there is 0.79 x 1.33 or 1.05 sq. in. of steel in the outer face. This is more than is required to resist the temperature stress, therefore, the wall will not crack.

N.6. DIGESTION TANK OPERATION

If the existing digestion tanks were cleaned and grit accumulation kept to a minimum, the following hydraulic and volatile matter loading conditions would prevail.

N.6.1. Hydraulic Residence Time

Maximum $$\frac{13{,}147{,}055 \text{ gallons of capacity}}{894{,}786 \text{ gallons of sludge per day}}$$

= 14.7 days

Average $$\frac{13{,}147{,}055 \text{ gallons of capacity}}{1{,}047{,}424 \text{ gallons of sludge per day}}$$

= 12.6 days

Minimum $$\frac{13{,}147{,}055 \text{ gallons of capacity}}{1{,}200{,}062 \text{ gallons of sludge per day}}$$

= 11.0 days

N.6.2. Volatile Matter Loading

Minimum

$$\frac{276{,}038 \text{ pounds volatile matter per day}}{1{,}757{,}628 \text{ cubic feet of capacity}}$$

$$= 0.157 \text{ pounds volatile matter per cubic foot per day}$$

Average

$$\frac{319{,}030 \text{ pounds volatile matter per day}}{1{,}757{,}628 \text{ cubic feet of capacity}}$$

$$= 0.182 \text{ pounds volatile matter per cubic foot per day}$$

Maximum

$$\frac{362{,}022 \text{ pounds volatile matter per day}}{1{,}757{,}628 \text{ cubic feet of capacity}}$$

$$= 0.206 \text{ pounds volatile matter per cubic foot per day}$$

N.7. GAS PRODUCTION

Gas production is expected to range from 12 to 14 cubic feet per pound of volatile matter reduced. Volatile matter reduction is expected to be 53.5 percent (same as mesophilic system but with a shorter hydraulic detention period).

Volatile matter reduced

$$(0.535)(319{,}030) = 170{,}681 \text{ pounds per day}$$

$$\frac{170{,}681 \text{ pounds}}{\text{day}} \times \frac{12 \text{ cubic feet}}{\text{pound}} = 2{,}048{,}172 \text{ cubic feet per day}$$

$$\frac{170{,}681 \text{ pounds}}{\text{day}} \times \frac{14 \text{ cubic feet}}{\text{pound}} = 2{,}389{,}534 \text{ cubic feet per day}$$

Gas consumption to meet winter and summer sludge heating requirements (based on 12 digestion tanks) would be:

Winter operation:

$$\frac{7.8 \times 10^{8} \text{ BTU}}{\text{day}} \times \frac{1 \text{ cubic foot gas}}{600 \text{ BTU}} \times \frac{1 \text{ BTU delivered}}{0.8 \text{ BTU utilized}}$$

$$= 1{,}625{,}000 \text{ cubic feet per day}$$

Summer operation:

$$\frac{5.4\ \text{x}10^{8}\ \text{BTU}}{\text{day}} \ \text{x}\ \frac{1\ \text{cubic foot gas}}{600\ \text{BTU}} \ \text{x}\ \frac{1\ \ \text{BTU delivered}}{0.8\ \text{BTU utilized}}$$

$$= \ 1{,}125{,}000\ \text{cubic feet per day}$$

Average expected excess:

$$\frac{(2{,}048{,}172 + 2{,}389{,}534)}{2}\ \frac{\text{cubic feet}}{\text{day}} - \frac{(1{,}625{,}000 + 1{,}125{,}000)}{2}\ \frac{\text{cubic feet}}{\text{day}}\ .$$

= 843,853 cubic feet per day available to be sold

APPENDIX O

SECTION 6.3 SUPPORTING CALCULATIONS

O.1. DIGESTED SLUDGE VACUUM FILTER OPERATION

Each vacuum filter has 600 square feet of filtering area. Assuming a minimum yield for this sludge, mesophilically and thermophilically digested, of 3.0 pounds per hour per square foot, each filter would be able to dewater:

$$\frac{3.0 \text{ pounds}}{\text{square foot-hour}} \times \frac{600 \text{ square feet}}{\text{filter}} \times \frac{24 \text{ hours}}{\text{day}}$$

$= 43{,}200$ pounds per day

As indicated elsewhere in this Appendix, if the maximum sludge production generated at 334 million gallons per day influent flow rate was mesophilically-thermophilically digested there would be 316,519 pounds per day to dewater. The maximum number of filters required would be:

$$\frac{316{,}519 \text{ pounds}}{\text{day}} \times \frac{1 \text{ filter-day}}{43{,}200 \text{ pounds}}$$

$= 7.3$, or 8 filters required

Assuming an average cake solids of 16.5 percent, the maximum number of wet tons to be hauled away including the polymer conditioner would be:

$$\left[\frac{316{,}519 \text{ pound sludge}}{\text{day}} + \frac{640 \text{ pounds polymer solids}}{\text{day}}\right] \times \frac{1 \text{ ton}}{2000 \text{ pounds}} \times \frac{1}{0.165}$$

$= 961$ wet tons per day maximum

O.2. EFFECT ON SYSTEMS HEATING CAPABILITIES DUE TO DUAL HEATING SYSTEM REQUIREMENTS

O.2.1. Mesophilic System Heat Requirements

The following calculations and numbers are derived and discussed in Appendix G.

Winter operation:

$$\left[\frac{1,148,161\ \text{BTU}}{\text{MGIF*}} \times \frac{334\ \text{MGIF}}{24\ \text{hours}}\right] + \frac{3,195,654\ \text{BTU}}{\text{hour}}$$

$= 19,174,228$ BTUs per hour

Summer operation:

$$\left[\frac{528,627\ \text{BTU}}{\text{MGIF}} \times \frac{334\ \text{MGIF}}{24\ \text{hours}}\right] + \frac{1,835,250\ \text{BTU}}{\text{hour}}$$

$= 9,191,976$ BTUs per hour

0.2.2. Thermophilic System Heat Requirements

Heat required for sludge both winter and summer:

$$\frac{1,200,062\ \text{gallons}}{\text{day}} \times \frac{8.64\ \text{pounds}}{\text{gallon}} \times \frac{(122-95)^\circ\text{F}}{1} \times \frac{1\ \text{day}}{24\ \text{hours}}$$

$= 11,664,603$ BTU per hour

Heat required for conductive/convective losses:

Roof - winter operation:

$$\frac{(0.23\ \text{BTU})(33,205\ \text{square feet})(122-0)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 931,732\ \text{BTU per hour}$$

Roof - summer operation:

$$\frac{(0.23\ \text{BTU})(33,205\ \text{square feet})(122-60)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 473,503\ \text{BTU per hour}$$

Walls - winter operation:

$$\frac{(0.13\ \text{BTU})(35,577\ \text{square feet})[(0.59)(122-0)^\circ\text{F} + (0.41)(122-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

$= 431,513$ BTU per hour

Walls - summer operation:

$$\frac{(0.13\ \text{BTU})(35,577\ \text{square feet})[(0.59)(122-60)^\circ\text{F} + (0.41)(122-70)^\circ\text{F}]}{\text{hour-square foot-}^\circ\text{F}}$$

$= 267,788$ BTU per hour

---

* MGIF = million gallons influent flow.

Floor - summer and winter operation:

$$\frac{(0.11\ \text{BTU})(45{,}195\ \text{square feet})(122-40)^\circ\text{F}}{\text{hour-square foot-}^\circ\text{F}} = 407{,}659\ \text{BTU per hour}$$

Sludge piping heat losses in thermophilic section are minimal.

Total hourly winter heat requirements:

$$\frac{11{,}664{,}603\ \text{BTU}}{\text{hour}} + \frac{1{,}772{,}186\ \text{BTU}}{\text{hour}} = \frac{13{,}436{,}789\ \text{BTU}}{\text{hour}}$$

0.3. DIGESTION TANK OPERATION

At the present time, the existing tank layout lends itself to using digesters 1 through 8 as the mesophilic units and digesters 9 through 12 as the thermophilic units. If all the existing digestion tanks were cleaned and grit accumulation kept to a minimum, the following hydraulic and volatile matter loading conditions would prevail.

0.3.1. Hydraulic Residence Time

	Mesophilic System Digesters 1-8	Thermophilic System Digesters 9-12
Maximum	$\frac{8{,}768{,}000\ \text{gallons of capacity}}{894{,}786\ \text{gallons of sluge per day}}$ = 9.8 days	$\frac{4{,}384{,}000\ \text{gallons of capacity}}{894{,}786\ \text{gallons of sludge per day}}$ = 4.9 days
Average	$\frac{8{,}768{,}000\ \text{gallons of capacity}}{1{,}047{,}424\ \text{gallons of sluge per day}}$ = 8.2 days	$\frac{4{,}384{,}000\ \text{gallons of capacity}}{1{,}047{,}424\ \text{gallons of sludge per day}}$ = 4.1 days
Minimum	$\frac{8{,}768{,}000\ \text{gallons of capacity}}{1{,}200{,}062\ \text{gallons of sluge per day}}$ = 7.2 days	$\frac{4{,}384{,}000\ \text{gallons of capacity}}{1{,}200{,}062\ \text{gallons of sludge per day}}$ = 3.6 days

Hydraulic residence times are too short for both a mesophilic and a thermophilic system. If two digesters of same size were added to both processes, the new hydraulic residence times would be:

	Mesophilic system 10 tanks	Thermophilic system 6 tanks
Maximum	12.2	7.3
Average	10.5	6.3
Minimum	9.1	5.5

0.3.2. Volatile Matter Loading

Volatile matter loading based on 10 mesophilic tanks only:

Minimum $\dfrac{276{,}038 \text{ pounds volatile matter per day}}{1{,}464{,}690 \text{ cubic feet of capacity}}$

= 0.19 pounds volatile matter per cubic foot per day

Average $\dfrac{319{,}030 \text{ pounds volatile matter per day}}{1{,}464{,}690 \text{ cubic feet of capacity}}$

= 0.22 pounds volatile matter per cubic foot per day

Maximum $\dfrac{362{,}022 \text{ pounds volatile matter per day}}{1{,}464{,}690 \text{ cubic feet of capacity}}$

= 0.25 pounds volatile matter per cubic foot per day

0.4. GAS PRODUCTION

Gas production is expected to range from 12 to 14 cubic feet per pound of volatile matter reduced. The volatile matter reduction should be a minimum of 60 percent.

Volatile matter reduced:

(0.60)(319,030) = 191,418 pounds per day

$$\frac{191{,}418 \text{ pounds}}{\text{day}} \times \frac{12 \text{ cubic feet}}{\text{pound}} = 2{,}297{,}016 \text{ cubic feet per day}$$

$$\frac{191{,}418 \text{ pounds}}{\text{day}} \times \frac{14 \text{ cubic feet}}{\text{pound}} = 2{,}679{,}852 \text{ cubic feet per day}$$

Gas consumption to meet winter and summer sludge heating requirements (based on 16 digestion tanks) would be:

Winter operation:

$$\frac{7.9 \times 10^8 \text{ BTU}}{\text{day}} \times \frac{1 \text{ cubic foot gas}}{600 \text{ BTU}} \times \frac{1 \text{ BTU delivered}}{0.8 \text{ BTU utilized}}$$

= 1,645,833 cubic feet per day

Summer operation:

$$\frac{5.3 \times 10^{8}\ \text{BTU}}{\text{day}} \times \frac{1\ \text{cubic foot gas}}{600\ \text{BTU}} \times \frac{1\ \text{BTU delivered}}{0.8\ \text{BTU utilized}}$$

$= 1{,}104{,}167$ cubic feet per day

Average expected excess:

$$\frac{(2{,}297{,}016 + 2{,}679{,}852)}{2}\frac{\text{cubic feet}}{\text{day}} - \frac{(1{,}654{,}833 + 1{,}104{,}167)}{2}\frac{\text{cubic feet}}{\text{day}}$$

$= 1{,}108{,}934$ cubic feet per day available to be sold.

APPENDIX P

THERMOPHILIC DIGESTION REFERENCES FOR SECTION 2.2 AND 6.2

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