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FEASIBILITY STUDY FOR ALTERNATE FUELS PRODUCTION: UNCONVENTIONAL NATURAL GAS FROM WASTEWATER TREATMENT PLANTS

Volume II Appendix D

December 1981

Acurex Project 7779
Contract No. DE-FG01-80RA50304

For
U.S. Department of Energy
Office of Procurement Operations
400 First St. N.W.
Washington, DC 20585

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FEASIBILITY STUDY
FOR ALTERNATIVE
FUELS PRODUCTION

FINAL REPORT

SEPTEMBER 30, 1981

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FEASIBILITY STUDY
FOR ALTERNATIVE
FUELS PRODUCTION

FEASIBILITY STUDY FOR ALTERNATE FUELS PRODUCTION

INTRODUCTION

From the turn of the century to about 1936, wastewater solids were settled and treated in a single dual purpose tank, called an Imhoff tank. Named after its German inventor, it was used extensively in Europe and later in the United States for settling solids (P.16). Separate sludge digestion did not start in the United States until about 1911 in Baltimore. It was successful and initiated 20 years of further development of the process. Since the main objective was sludge stabilization, the studies determined design criteria for process control. This included; sludge heating, digester temperature, pH and alkalinity. Extensive lab and pilot plant tests were made between 1910 and 1940 applying these environmental factors to the anaerobic digestion process. In an effort to improve sludge stabilization, the volume of methane gas produced was increased.

The discovery of methane gas as a by-product of sludge stabilization was not new. Donald Cameron first recognized this combustible gas and its potential as an energy source in 1895. But the cost of energy was very low then and there was no incentive to use the digester gas. Much of it was flared as a waste product.

The biochemical reactions of anaerobic digestion occupied much of the research during the late 1940's and 1950's. Operational control of contact time between the food (organics) and the bacteria became important. Narrower ranges of temperature and pH were determined to keep the methane forming bacteria active. These bacteria were considered most critical in producing a well stabilized sludge.

With the national economic boom of the sixties, new methods of sludge treatment were studied and interest in anaerobic digestion declined. Energy intense alternatives were developed. But fossil fuels soon became scarce and expensive and researchers returned to the anaerobic digestion process to investigate the resourcefulness of the digestion gas by-product discovered nearly 100 years before.

The increasing concern for energy conservation continues into the 1980's; and research to optimize and purify digester gas is being developed as a fossil fuel alternative.

PURPOSE AND SCOPE

The U.S. Department of Energy (DOE) has been concerned with energy waste at wastewater treatment plants by large amounts of unused digester gas. It is evident that utilization of this gas would result in savings of public monies. Consequently, the DOE awarded a contract to the Acurex Corporation to explore both the maximizing of production and utilization of digester gas.

The portion of this study subcontracted to John Carollo Engineers can be outlined in three basic tasks.

TASK 1. OPTIMIZATION OF DIGESTER GAS PRODUCTION

This task included a detailed literature search to develop the practical and proven techniques for optimizing methane gas production from anaerobic digesters. It included a description of the impact of the following parameters on increasing gas production:

1. Solids retention time;
2. Thickening of sludge prior to digestion;

3. Mixing of the sludge;
4. Operating temperature in the digester;
5. Digester heating methods; and
6. Insulation of digester walls and roof.

The goal was to select the best techniques from a comparative economical and technological evaluation for new and existing wastewater plants.

TASK 2. DIGESTER GAS SCRUBBING PROCESSES

This task included a literature review of the viable digester gas scrubbing systems which can remove hydrogen sulfide and carbon dioxide

Sufficient information was to be included to permit an economical evaluation of the best practical scrubbing system for application to digester gas.

TASK 3. APPLICATION TO THE EAST BAY MUNICIPAL UTILITY DISTRICT WATER POLLUTION CONTROL PLANT

The alternatives selected from Task 1 were to be compared with the present performance of the digester gas production at the EBMUD Wastewater Treatment Plant. Where applicable, design criteria and costs were included, along with a schematic and plant layout.

CHAPTER I

PROCESS SURVEY

ANAEROBIC DIGESTION

Figure 1 is used as an example of a typical biological secondary Wastewater Treatment Plant. Sludge characteristics from both primary settled sludge and the waste biological sludge from secondary settling determine the design parameters of the sludge handling processes.

The sludge handling processes, as identified in Figure 2, include: thickening, stabilization, conditioning, dewatering, drying and final disposal. For purposes of this report sludge stabilization will include only anaerobic digestion. The various sludges fed into the anaerobic digester are presented in Figure 3. Sludge can be delivered to the digester in three forms. The waste biological sludge can be thickened and combined with the raw primary sludge, as in Figure 3A. The sludges can be combined with or without thickening, as in Figure 3B. Or there can be mechanical or gravity thickening of the primary sludge combined with separate thickening of the waste biological sludge, as in Figure 3C.

Anaerobic Digestion Process. Anaerobic digestion is a biological process whereby bacteria consume organic matter in the sludges to produce carbon dioxide and methane. Optimizing methane production consists of keeping the bacteria active and maintaining a food supply as large as possible. The specific variables involved will be discussed later.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
FLOW SCHEMATIC
SECONDARY WASTEWATER TREATMENT PLANT

FIGURE 1

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
FLOW SCHEMATIC
SOLIDS HANDLING PROCESSES

FIGURE 2

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
FLOW SCHEMATIC
THICKENING / STABILIZATION

FIGURE 3

The digestion process occurs in the absence of free oxygen. It is a fermentative process. That means that although new cells are produced as the organics in the sludge are consumed, the growth of the bacteria is relatively small. Thus, there is an overall reduction in the sludge volume that must be handled for final disposal.

There are two loading methods for anaerobic digestion. The standard rate (conventional) design consists of an unheated circular tank with either a fixed or floating cover. (Figure 4A) (B.27) Sludge is usually pumped into the middle of the tank. The sludge separates into layers of gas, scum, supernatant, an active layer and a layer of stabilized solids which are removed at the tank bottom. The digesters are called standard rate because of their low feed rate of volatile (organic) solids; 0.04 to 0.10 pounds of volatile solids per cubic feet per day ($\text{lb VS/ft}^3/\text{day}$). (A.11)

High rate digestion occurs in a mixed tank. (Figure 4B) The more complete the mixing, that is, the larger the active zone, the better the stabilization since the microorganisms have better contact with the food. Again the covers can be either fixed or floating. Sludge stabilization occurs in less time and at higher feed rates; 0.15 to 0.40 $\text{lbs VS/ ft}^3/\text{day}$.

For sewage plants treating flows greater than 1 MGD (million gallons per day), a combination of the two digesters is common. This is called two-stage digestion. The first stage (primary digester) is generally a mixed, high-rate digester. The second stage (secondary digester) is an unmixed, standard-rate unit. Its purpose is to store and concentrate the sludge prior to final disposal. Unfortunately, good

A. STANDARD (Conventional) RATE

B. HIGH RATE

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

ANAEROBIC DIGESTERS

FIGURE 4

separation between the solids, liquid and gas does not always occur. Table 1 (P.16) summarizes the differences in the two basic types of anaerobic digesters.

Regardless of the mode of operation, the basic biochemical reactions are the same:

The microorganisms responsible for the decomposition of the complex organics into the organic acids are the acid-forming bacteria. These consist of both facultative and anaerobic bacteria which hydrolyze and ferment the organics to simple organic acids, such as acetic and propionic.

These simple organic acids are then reduced by methane-forming bacteria to methane, carbon dioxide and smaller amounts of hydrogen sulfide, hydrogen, and nitrogen. Research has shown that the methane fermentation from the acids is accomplished in about 5 days.

The two reactions occur simultaneously and an equilibrium must be maintained if gas production is to be optimized. The methane-forming bacteria have a slower growth rate than the acid-forming bacteria and are very sensitive to pH. They do not function below 6.2 and begin dying off at a pH above 7.6. (P.16)

Wastewater data shows that the amount of methane that can be produced per pound of food utilized by the bacteria is between 12 and 18 cubic feet of gas per pound of volatile solids destroyed. (B.9)

Table 1

ANAEROBIC DIGESTION LOADING

ITEM	STANDARD RATE	HIGH RATE
Heated	No	Yes
Loading (lb. VS/ft ³ /day)	0.04-0.10	0.15-0.40
Sludge feeding and withdrawal	Intermittent	Continuous
Detention (days)	30 or more	15 or less
VS Reduction (1) (%)	50-60	50-60
Purpose	Sludge separation	Sludge mixing and stabilization

(1) VS reduction is dependent on whether the incoming sludge is combined, separated or thickened; and the initial VS content.

DIGESTER GAS

Digester gas characteristics have been identified as generally containing 65 percent methane (CH_4), 30 percent carbon dioxide (CO_2) and 5 percent of other gases including; hydrogen sulfide (H_2S), nitrogen (N_2) and hydrogen (H_2). (A.13), (P.16).

The gas production is particularly dependent on digester mixing and sludge temperature. (A.13) Since gas production and sludge stabilization are directly related, improved stabilization should increase gas production. Because the methane-forming bacteria have a slower growth rate than the acid-formers, during the early stages of digestion (e.g. plant start-up) more carbon dioxide and hydrogen sulfide will be produced than methane. Methane production will slowly increase. When it reaches 56 percent of the digester gas volume, it can be burned. But the gas is not usable as a fuel until the methane content is, at least, 60 percent. (P.14) The numbers were generated by wastewater plant operators for keeping the heat value of the digester gas at above 500 BTU per cubic foot of gas. The range of methane in the digester gas can be from 40 to 75 percent. A review of actual Wastewater Treatment Plant operating data narrows this range to 60 to 70 percent with carbon dioxide between 25 and 30 percent and the remaining 5 to 10 percent consisting of hydrogen, nitrogen and hydrogen sulfide. Data has also shown that the methane content from a particular plant varies less than 5 percent under normal operating conditions. This indicates that an attempt to increase the percent of methane in the digester gas is not as realistic as increasing the total digester gas produced. (L.19)

The usefulness of digester gas is measured in terms of its heating potential or heat value in Btu (British Thermal Units). A Btu

is defined as the heat required to raise one pound of water one degree fahrenheit. (P.19) A simple mathematic conversion of Btu provides electrical energy potential. (2.928×10^{-4} KWH/Btu, KWH is a kilowatt-hour). The heat value for digester gas ranges from 550 to 650 Btu/ft³ of gas. Some New York City plants report as much as 654 Btu available from a gas with 64.4 percent methane. Other plants report 590 to 600 Btu. These latter values are more typical. In no instances were values greater than 650 obtained on a routine basis (L.9).

ENERGY BALANCE

To realize the significance for optimizing digester gas production, it is important to understand the methods and quantities of energy used in a Wastewater Treatment Plant. Daily energy consumption by the various unit processes shown in Figure 1 are provided in Table 2 (0.1). These numbers were generated by R. Smith in 1973. They are average numbers and are provided for comparison of activated sludge treatment with trickling filters and primary treatment. (Actual plant energy consumption varies greatly.) The headworks includes; bar screens, comminutors and grit removal. Thickening and dewatering was not included because it was not representative of a typical wastewater treatment plant in 1973. More specific energy requirements for operating sludge handling equipment is provided in Chapter IV.

Table 2 shows that about 8200 BTU/capita/day (based on a flow of 100 gallons of wastewater per capita per day) is required to operate an activated sludge plant. This does not consider that a plant's energy can be satisfied by using oil, coal, natural gas, hydroelectric power et. al. It assumes all of the kilowatts needed to be generated by electricity and converted at the rate of 3415 BTU per KWH.

TABLE 2

TYPICAL DAILY ENERGY USAGE
FOR VARIOUS UNIT OPERATIONS

(KWH, UNLESS OTHERWISE NOTED)

PROCESS	PRIMARY			TRICKLING FILTER			ACTIVATED SLUDGE		
	1 MGD	10 MGD	100 MGD	1 MGD	10 MGD	100 MGD	1 MGD	10 MGD	100 MGD
Headworks	18.2	65.9	245	18.2	65.9	245	18.2	65.9	245
Primary Sedimentation	31	120	730	31	120	730	31	120	730
Secondary Treatment	-	-	-	180	1,700	16,000	580	5,700	56,000
Secondary Sedimentation	-	-	-	31	120	730	31	120	730
Sludge Pumping	0.6	6.4	64	2.7	27	270	2.7	27	270
Miscellaneous Pumping	150	1,500	13,000	150	1,500	13,000	150	1,500	13,000
Chlorination	0.7	82	830	0.7	0.7	270	0.7	0.7	270
Digester Mixing and Pumping	100	340	1,500	120	470	1,900	120	470	1,900
Lights and Miscellaneous	57	210	2,400	57	210	2,400	57	210	2,400
Total Electrical Use	360	2,300	19,000	600	42,000	36,000	1000	8,200	76,000
Total Usage in BTU @ 3413 BTU/KWH (x 10 ⁶)	1.2	7.9	64.9	2.0	14.3	122.9	3.4	28.0	259.5
Digester Heating Requirement (x 10 ⁶ BTU)	3.1	31	310	5.4	54	540	5.0	54	540
Total Energy Requirement (x 10 ⁶ BTU)	4.3	38.9	374.9	7.4	68.3	662.9	8.4	82.0	799.5

This amount of energy represents only several tenths of a percent of the total per capita energy used in this Country. Although this amount seems small, the total annual energy use for all wastewater treatment plants in this country is about 10^{14} Btu. This is roughly the equivalent amount of energy in 20 million barrels of crude oil. (0.19)

Since digester gas has a potentially comparative heat value with other fuels (see Table 3), it should not be discarded as a fuel source for meeting some of the wastewater treatment plant energy needs.

To satisfy an energy balance, the energy produced in wastewater treatment plant operations must be evaluated and compared with the energy demands. Wesner and Clarke (L.19) have generated some gas production data based on:

1. An activated sludge plant with 60 percent volatile solids in the primary sludge, 80 percent volatile solids in the waste activated sludge (WAS) and 50 percent volatile solids destroyed in digestion;
2. 15 cubic feet³ of gas produced per lb of volatile destroyed; and
3. 600 Btu per cubic feet³ of gas.

Some of this data is presented in Table 4. Assuming 7000 Btu/HP-hr for internal combustion engines (@ 36.4% efficiency) and 11,400 Btu/KW-hr for engine-generators (@ 30.0% efficiency), the heat available (Btu) can be converted to an equivalent amount of electrical energy (KWH). Similar data generated by James Smith in 1976 was about 3 percent lower. He used lower volatile solids percentages in the primary and waste activated sludges. (D.10)

TABLE 3
HEAT VALUE OF FUELS

Fuel Oil	140,000 Btu/gal of fuel (No. 2) 155,000 Btu/gal of fuel (No. 6)
Coal Fuel (bituminous)	13,500 Btu/lb of coal
Natural gas	1,050 Btu/ft ³
Manufactured gas	
a. Carbonated water or coke gas	550 Btu/ft ³
b. Butane	2400 Btu/ft ³
Digester Gas	590-650 Btu/ft ³

A direct comparison can be made between the heat available from digester gas, Table 4, and the total energy needs of an activated sludge plant, Table 2. About one-half of the total energy needs of an activated sludge plant can be provided from digester gas; regardless of the size of the plant.

An additional amount of energy demand can be satisfied with heat recovery; for example, from gas-driven engines. This extra heat energy can provide as much as 45 percent more energy.

ON-SITE GAS USES

Digester gas can be used to provide heat and power for other Wastewater Treatment Plant operations, as previously mentioned. This includes; building heat, incineration facilities, internal combustion (IC) engines to operate electric generators, pumps, lights, air compressors, etc. Additionally, power can be provided to run the blowers for aeration. Since the amount of gas produced is not uniform, and has no relationship to usage, storage should be considered. Storage capacity is recommended at 25 to 50 percent of the average daily gas produced. (P.22)

The Federal Environment Protection Agency estimates that a plant must treat, at least, 10 million gallons per day to benefit, economically, from on-site power generation. (D.16)

OFF-SITE GAS USES

Economically marketing digester gas requires its heat value to be competitive with other fuels as shown in Table 3; or electricity can be generated and sold to off-site customers. In either case, additional costs are expended in using the digester gas.

TABLE 4

DIGESTER GAS PRODUCTION

WWTP Capacity (MGD)	Total Dry Solids to digester (lbs/day)	VS destroyed (lbs/day)	Gas Produced (Ft ³ /day)	Heat Available (x10 ⁶) (BTU/day) (KWH/yr)	
1	2,100	723	10,845	6.5	0.21
5	10,500	3,615	54,225	32.5	1.04
10	21,000	7,230	108,450	65.0	2.08
25	52,500	18,025	271,125	162.5	5.2
50	105,000	36,105	542,250	325.0	10.4
75	157,500	54,225	813,375	487.5	15.6
100	210,000	72,300	1,084,500	650.0	20.8

Since gas production is not consistent, low pressure storage is typically provided so that a continuous stream of gas is available for generating electricity. The facility could be powered by reciprocating engines, steam turbines or gas turbines. Dual fueled engines have operated successfully at wastewater treatment plants for some 20 years. These engines use a combination of digester gas and fuel oil for operation. Gas driven turbines and engines have had a history of maintenance problems due to the contaminants found in digester gas; specifically, hydrogen sulfide. The problem is solvable. One approach is to remove the hydrogen sulfide prior to its use in engines. Another approach is to operate the engines at a higher temperature (greater than 200°F), above the dew point to prevent condensation of the hydrogen sulfide.

Some manufacturers recommend using a low grade fuel, such as digester gas, in a turbocharged engine. In this case, the gas has to be fed at higher pressures through a smaller orifice. Consequently, costs may be higher and the gas impurities may cause problems in the flow through the smaller orifice.

In order for digester gas to meet the requirements of natural gas pipeline quality, additional treatment is needed to lower both the moisture and carbon dioxide levels. Unprocessed digester gas is generally saturated with water and contains nearly 30 percent of carbon dioxide.

Additional uses of digester gas include combustion of the gas and the sale of the resulting heat in the form of steam or hot water. Sale of compressed gas as a vehicle fuel is also possible.

GAS PRODUCTION VARIABLES

The key to successful operation of an anaerobic digester, and optimization of the gas produced, is to maintain proper environmental conditions so that a balance exists between the acid-forming and methane-forming bacteria. Imbalances result in poor volatile solids destruction and low gas production.

Operating a digester is complicated and requires a high level of skill to achieve success. The variables considered to be most important for optimizing methane gas production include:

1. Sludge Thickening - provides more volatile solids per unit volume in the digester;
2. Solids Retention Time - directly related to the volatile solids reduction in the digester;
3. Digester Mixing - provides the contact between micro-organisms and volatile solids;
4. Sludge Temperature - an optimum operating temperature exists in both the mesophilic and thermophilic ranges at which digester gas production is optimized;
5. Digester Heating - provides heat to the incoming sludge and maintains the sludge temperature inside the digester; and
6. Insulation - a tool for energy conservation rather than gas production as it reduces heat losses in the digester system.

Each of the variables is addressed in detail in the following Chapter.

CHAPTER II

VARIABLES AFFECTING DIGESTER GAS PRODUCTION

INTRODUCTION

The amount of gas produced from anaerobic digestion depends on the amount of organics consumed in the process. This may be affected by the quantity of sludge fed to the digester, the volatile solids content of this untreated sludge, and the operating characteristics of the digester.

Considering the treatment processes of Figure 1, biological secondary treatment is usually accomplished by activated sludge, trickling filters or rotating biological discs. The sludges generated are similar in quality but vary in quantity as indicated in Table 5. These numbers are an average from those reported by Fair and Imhoff, Keefer and Eckenfelder and McCabe. (A.10)

SLUDGE THICKENING

Sludge thickening can produce a sludge with the same volatile solids content (percent) as the raw sludge but at a reduced volume. Although the total solids content (percent) will be increased, the fraction of the total solids which consist of the volatile solids, does not change. This can result in cost savings for digester tanks in a new plant or in gaining detention time in a digester by adding thickening to upgrade an existing plant. Thickening sludge solids should not be confused with dewatering. The latter process occurs after digestion for concentrating solids to greater than 15 percent. At this solids concentration, positive displacement pumps are required.

Thickening can be effectively accomplished with gravity thickeners, air flotation thickeners and centrifuges.

TABLE 5

TYPICAL QUANTITIES OF BIOLOGICAL
SLUDGES PRODUCED IN SECONDARY TREATMENT

Treatment Process	Sludge Production (lb. dry solids/MG)
Primary	1240
Primary + Trickling Filter	1700
Primary + Activated Sludge	2700

GRAVITY THICKENING. Either primary or a combined primary and waste biological sludge can be effectively thickened with gravity thickeners. Table 6 (P.16) lists the solids concentrations for various gravity thickened sludges, both separate and combined. A mechanical gravity thickener is generally used for combined sludges. Figure 5 is a typical example of a mechanical gravity thickener. Excessive grit may accumulate in the thickened sludge. This could affect pumping and, because grit is non-volatile, reduce the gas production potential of the sludge. The grit can accumulate in the digester also affecting capacity and detention time. Consequently, grit removal in the Wastewater Treatment Plant headworks must be effective and efficiently maintained.

As can be seen in Table 6, gravity thickeners are not very effective for thickening separate activated sludge. These numbers by Metcalf and Eddy are in agreement with the Federal Environmental Protection Agency's claim that a maximum of only 3 percent solids can be achieved on activated sludges.

The primary advantage of gravity thickening is its low operation and maintenance costs, particularly power consumption.

AIR FLOTATION THICKENING. Flotation thickening uses rising gas bubbles, usually air, to increase the bouyancy of solid particles and force them to separate from the water and rise to the surface. Chemicals are often added to aid in solids agglomeration and settling. Air flotation thickening is best for biological sludges with less than a 2 percent solids concentration. Thickening to 4 or 5 percent can be easily achieved. Loading rates for dissolved air floatation (DAF) thickeners are given in Table 7 (A.12). Some plant data has shown these rates to be conservative and higher percent solids are achieved at or above these values.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**MECHANICAL GRAVITY
THICKENER**

FIGURE 5

TABLE 6
CONCENTRATIONS OF UNTHICKENED AND THICKENED
SLUDGES AND SOLIDS LOADINGS FOR GRAVITY THICKENERS

TYPE OF SLUDGE	SLUDGE SOLIDS (percent)		SOLIDS LOADING (lb/sq. ft./day)
	UNTHICKENED	THICKENED	
SEPARATE SLUDGES:			
Primary	2.5 - 5.5	8 - 10	20 - 30
Trickling Filter	4 - 7	7 - 9	8 - 10
Activated	0.5 - 1.2	2.5 - 3.3	4 - 8 (1)
COMBINED SLUDGES:			
Primary & Trickling filter	3 - 6	7 - 9	12 - 20 (2)
Primary & Activated	2.6 - 4.8	4.6 - 9.0	8 - 16 (3)

- (1) EPA recommends a 5-6 loading rate.
- (2) EPA recommends a 10-12 loading rate.
- (3) EPA recommends a 6-10 loading rate.

TABLE 7

SOLIDS LOADING RATES FOR DAF THICKENERS
TO PRODUCE A 4 PERCENT (MIN.) SOLIDS CONCENTRATION

TYPE OF SLUDGE	SOLIDS LOADING RATE (lbs/sq. ft./day)	
	with no chemical addition	with optimum chemical addition
Primary only	0.85 - 1.25	up to 2.5
Waste Activated Sludge	0.42	up to 2.0
Trickling Filter	0.6 - 0.8	up to 2.0
Primary + Waste Activated Sludge	0.6 - 1.25	up to 2.5

Figure 6 (A.10) is a typical schematic of a DAF thickener. The components usually consists of: a sludge feed pump and meter; polymer mixing and feeding equipment; flotation thickener; electrical control panel; air compressor and receiver; and thickened sludge recirculation pump. The main advantages of air flotation over gravity thickening of biological sludges include: better reliability; a thicker product; and less odor problems. The chief disadvantages are: higher operating costs and little storage capacity.

CENTRIFUGAL THICKENING. There has been limited use of centrifugal thickening because of its substantial maintenance and high power costs. There are three general types: solid bowl, disc-nozzle and basket. The theory in each is basically the same. There is an acceleration of sedimentation through centrifugal force. Solids sink to the bottom and the liquid remains on the top.

Figure 7 (A.10) is a schematic of a solid bowl centrifuge. Although 7 percent solids can be achieved, costs for chemicals can be high. Advantages include: easy installation; quiet and clean operation; and a low capital cost as compared to other centrifuge units. But it is a potentially high maintenance item requiring specially skilled operators. Present designs are for handling waste biological sludges only. A study in 1978 at the Village Creek Wastewater Treatment Plant in Fort Worth, Texas showed a high percent solids concentration without chemical addition (A.13).

Centrifuge thickening at the East Bay Municipal Utility District Wastewater Treatment Plant in Oakland, California, had produced a high level of ammonia in the thickened sludge. Consequently, toxicity

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

DISSOLVED AIR FLOTATION THICKENER

FIGURE 6

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
SOLID BOWL CENTRIFUGE

FIGURE 7

occurred in the digesters to the microorganisms. Such problems must be considered when selecting thickeners which produce solids greater than 7 percent. (Ammonia toxicity is discussed in more detail at the end of this Chapter.)

A schematic of a disc-nozzle centrifuge is shown in Figure 8 (A.11). This method of centrifugal thickening dates as far back as 1937. (A.13) The system was used for industrial wastes and has not, as yet, been fully redesigned for municipal sludge application. Pilot studies at Fort Worth, Kelowna, (British Columbia) and Idaho Falls showed the disc-nozzle to fall behind in comparison with other thickeners. With additional modifications to alleviate plugging, Fort Worth has installed disc-nozzles with limited success. The chief advantage of the disc-nozzle centrifuge is its large solids handling capacity within a very small space. The disadvantages include: extensive prescreening for grit removal; high maintenance; and the requirement of specially skilled personnel. Capital costs are similar to the solid bowl centrifuge.

The third type of centrifugal thickener is the imperforate basket shown in Figure 9 (A.11). Extensive field testing has consistently produced a sludge with 9 to 10 percent solids. But few full-scale operations use basket thickening. The latest is in Los Angeles County. The basket centrifuge can be used for thickening or dewatering. It is not as sensitive to grit as the disc-nozzle and has the lowest operation and maintenance costs of the three. The disadvantages include: high capital cost; special structural support; and the fact that feed and discharge is intermittent to the digester.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

DISC - NOZZLE CENTRIFUGE

FIGURE 8

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**IMPERFORATE BASKET
CENTRIFUGE**

FIGURE 9

Cost benefits for sludge thickening are indicated in Figure 10. As the percent solids in the sludge fed to the digester increases, the required digester volume decreases. Hence, there is a savings in costs to construct the digester.

MISCELLANEOUS THICKENING METHODS. Elutriation basins have had limited satisfactory use for washing and thickening digested primary sludges. Elutriating combined sludges require a chemical for flocculation.

Ultrafiltration of waste biological sludges has also met with limited success. Power requirements are relatively high for obtaining 6 percent solids.

Digested sludge lagoons have been used successfully but it is often too expensive because of the required large amounts of land.

The anaerobic digester can be used for thickening. At present, the process is not very reliable but the possibility of increasing methane production exists. No data has proven this method of thickening successful.

AMMONIA TOXICITY

The effects of ammonia-nitrogen on the anaerobic digestion process have been studied by McCarty and McKinney (A.12) and Albertson (B.1). Results of their work is presented in Table 8.

Ammonia gas (NH_3) is produced by the degradation of proteins and urea and can reach toxic levels in highly concentrated sludges. The reaction is pH dependent. At high pH levels, about 7.4, production of NH_3 gas is favored over the NH_4^- -ion. If the concentration of NH_3 reaches 1300 mg/l, toxicity may develop in the digester. If the NH_3

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
EFFECTS OF THICKENING
ON REQUIRED DIGESTION CAPACITY

FIGURE 10

TABLE 8
AMMONIA-NITROGEN EFFECTS ON ANAEROBIC DIGESTION

$\text{NH}_3\text{-N}$ (mg/l as N)	Effect
50-200	Beneficial
200-1,000	No adverse effects
1,500-3,000	Inhibitory at pH over 7.4
Above 3,000	Toxic

concentration is above 3000 mg/l, NH_4 -ion concentration will increase and cause toxicity. One solution is to dilute the incoming sludge to the digester.

When evaluating sludge thickening, consideration for NH_3 build-up should not be overlooked. Centrifuge thickening at the East Bay Municipal Utility District Wastewater Treatment Plant produced a sludge with 8 percent solids and NH_3 concentrations in the toxic range.

Separate digestion of primary sludge can produce a supernatant high in ammonia concentration. Return of this liquid to the plant head-works can cause toxicity in other biological treatment units.

Toxicity is also affected by the solids loading rate on the digester. Low rate digesters generally adjust easier to toxic shocks. Detention times are long enough to allow the organisms to acclimate themselves to the new environment.

SOLIDS RETENTION TIME

Factors influencing the solids retention time (SRT) include the volatile solids loading on the digester and the percent of volatile solids (volatile material) in the raw sludge. The latter relationship is expressed graphically in Figure 11 (P.22).

A review of wastewater treatment plant operation data by Woods and Malina (B.28) reveal that for a two-stage, high rate digester, 57 percent of the volatile solids are converted to a liquid or gas in the primary digester. An additional 2.8 percent are reduced in the secondary digester.

Whether the digester has low rate or high rate loading also affects the SRT. Low rate digestion requires between 30 and 60 days of detention. High rate loading requires between 10 and 20 days.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
VOLATILE SOLIDS REDUCTION
vs **SOLIDS RETENTION TIME**

FIGURE II

During gas production, the regeneration rate for the methane formers has been identified at about 10 days at 95 degrees F. (A.10) This can be considered the minimum SRT. McCarty recommends using a safety factor of 2.5 for a design SRT to maintain the growth of these methane forming bacteria. He uses the simple equation:

$$\frac{\text{SRT design}}{\text{SRT min}} = \text{Safety Factor}$$

The SRT is based on retention of the microorganisms and should not be confused with the hydraulic retention time (HRT). This is the retention time of the total liquid sludge. The SRT and HRT approach equality in a completely mixed system with no recycling.

A pilot study by O'Rourke at Stanford University showed that methane production varied with SRT as presented in Figure 12. (H.3) This was for high rate loading. There is no significant gain in methane production after 20 days SRT. Although the rate of methane production was very high in the first 10 days, the gas production activity made separation of the liquid and solids difficult. Therefore, additional days were needed. As the activity leveled off, separation and compaction occurred. This process would occur in a separate tank in two-stage digestion. Laboratory calculations based on process kinetics predict a critical SRT of 4.2 days at 95 degree F (35 degrees C) at which the digestion process will fail and the microorganisms will be washed out. (This should not be confused with the minimum SRT associated with the regeneration of methane-forming organisms.) Torpey's pilot-plant study of critical SRT resulted in 2.6 days at 99 degrees F. (0.21)

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
METHANE PRODUCTION
vs **SOLIDS RETENTION TIME**

FIGURE 12

Additional studies by Sawyer, Dague, and Walker (E.2, B.6, B.24) recommend 10 days minimum SRT for high rate digestion at temperatures between 95-100 degrees F. This corresponds to the minimum SRT for the regeneration of the methane formers. Similar studies for low rate loading were not available.

Temperature affects SRT in terms of gas production as shown in Figure 13 (A.11). The shape of the curves and the point at which they level out are strongly influenced by the temperature. At longer SRTs, more common in low rate digester loading, the effect of temperature is less pronounced.

In summary, then, low rate digestion requires at least 30 days SRT in the mesophilic temperature range to optimize gas production. Small variations in temperature and improved volatile solids reduction are not significant at higher SRTs.

High rate digestion requires about 20 days SRT in the mesophilic temperature range to optimize gas production. Digestion is influenced by temperature variances and should be kept between 95-99 (35 to 40 degrees centigrade) degrees F. Below 10 days SRT, the methane formers will begin to decline.

Other temperature impacts, including operating in thermophilic ranges will be discussed in the following section.

TEMPERATURE

Satisfactory digestion can occur at temperatures below 110 degrees F (the mesophilic range) and above 110 degrees F (the thermophilic range). The relationship between temperature and time (days of digestion) is shown in Table 9 which is a summary of data provided by

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**METHANE PRODUCTION vs TEMPERATURE
AND SOLIDS RETENTION TIME**

FIGURE 13

TABLE 9

TIME AND TEMPERATURE OF ANAEROBIC DIGESTION

	MESOPHILIC						THERMOPHILIC			
Temperature (degrees Fahrenheit)	50	60	70	80	90	100	110	120	130	140
Solids Retention Time (Days)	75	56	42	30	25	24	26	16	14	18

Fair and Moore (P.22) for primary sludge digestion and 55 percent volatile solids reduction. Optimum gas production in the mesophilic range was previously narrowed to 95 to 99 degrees F. Figure 14 (A.11) shows that gas production is optimized in the thermophilic range at about 134 degrees F, and in the mesophilic range the optimum temperature is narrowed to 96 degrees F. This graph is based on 23 separate studies by Schwerin at Marquette University in 1976. (H.4)

The City of Los Angeles - Hyperion Wastewater Treatment Plant has done extensive studying of thermophilic digestion. (J.5) One study concluded that nearly 17 percent more gas can be produced per day from thermophilic digestion. But there are disadvantages. More heat energy is needed. It can be as much as 69 percent more than the heat needed in mesophilic digestion. There is the possibility of solubilizing some heavy metals in the return supernatant. The thermophilic bacteria are more sensitive to pH and temperature variations. The advantages are that the digester sludge is more stable and dewaterers better.

Pilot studies by A. G. Fisher compared the amount of digester gas produced at 90 degrees F and 130 degrees F. Overall, there was little difference. Consequently, the benefit of thermophilic digestion seems more applicable to improving dewatering rather than optimizing gas production. (J.3)

Recently, a study by Environmental Technology Consultants, Inc. at the Rockaway, N.Y. Wastewater Treatment Plant attempted to reduce the volume of digested sludge produced for final disposal. Sludge digestion consisted of both mesophilic and thermophilic digesters

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
GAS PRODUCTION
vs **TEMPERATURE**

FIGURE 14

operating in series. Over 55 percent of the volatile solids were reduced in digestion, about 35 percent of the volatile solids were recycled and the remaining 10 percent were left in the sludge for final disposal. (0.14)

Considering this, a review of data from thirty wastewater treatment plants of varying sizes was made. It showed that the volatile solids reduction in mesophilic digestion averaged just over 47 percent. Consequently, the advantage in the combination mesophilic/thermophilic digestion seems to be that a smaller sludge volume is produced for final disposal with good dewatering characteristics but with only a slight increase in volatile solids destruction; hence, only a slight, if any, increase in digester gas production.

Accurate temperature readings are important to maintain bacterial growth. A thermometer can be installed in the recirculating sludge line from the digester to the heat exchanger. The temperature can also be measured from the flowing supernatant or from the heat exchanger inlet line. Regardless, the temperature should be kept constant at all levels within the digestion tank to avoid short circuiting. Mixing helps to accomplish this and is covered in the next section. Since the methane-forming bacteria are easily upset by temperature changes, variations throughout the tank should be no more than 1 or 2 degrees.

MIXING

Mixing is one of the more important controls for optimizing digester gas production. Good digester mixing will provide:

1. Immediate contact between the raw sludge and the active biomass (microorganisms);

2. Uniform distribution of food, microorganisms, alkalinity, heat, and waste bacterial products in the digester creating homogenous contents;
3. Utilization of as much of the total contents of the digester as possible;
4. Prevention of a surface scum layer and bottom deposition of grit and inert solids; and
5. Rapid distribution of toxic materials, minimizing their inhibitory effects on bacterial activity.

Mixing can be provided by:

1. Releasing compressed digester gas through diffusers near the digester bottom;
2. Mechanical mixing or pumping of sludge within the digester; and
3. Recirculating sludge with an external pump.

Although a certain amount of natural mixing occurs by rising gas bubbles in the sludge, artificial methods of mixing are essential to achieve effective digestion. Low rate digesters are usually not mixed. Consequently, only about 50 percent of the volume undergoes digestion. This digestion is partially attributed to the natural gas evolution and convection currents caused by temperature gradients. (E.4) Some reports show that the effective digestion volume of unmixed tanks is only 30 percent. (F.2) Regardless of this percentage, one thing is certain, the effectiveness of digestion in low rate, unmixed, digesters is very low.

The importance of mixing is amplified when thickening is provided. The feed solids concentration increases and the associated

higher viscosity tends to decrease the mixing effects of natural gas evolution and convectional currents. (E.2)

GAS MIXING. Gas mixing is one of the more common methods of digester sludge mixing. Generally, the gas is pulled from the tank, compressed and discharged through gas outlets or orifices inside the digester. The gas rises to the surface carrying sludge and creating a gas lift with a rolling action as shown in Figures 15A through D. Gas mixing can be accomplished as follows:

1. Injection of large gas bubbles at the tank bottom through large tube(s) (generally 6 to 12 inches in diameter) to create a piston-like pumping action and periodic surface agitation. (Figure 15A)

This method, although requiring low power provides a low level of mixing. It controls scum better than the other gas mixing systems.

2. Injection of gas bubbles through a series of lances suspended from the digester cover to as great a depth as possible, depending on cover travel. (Figure 15B.)

This method reduces solids accumulation on the tank bottom. Cleaning can be accomplished without sludge drawdown but clogging of the lances can occur.

3. Unconfined release of gas bubbles from spargers mounted on the digester floor. (Figure 15C)

There is no pumping action at the tank bottom. Consequently, solids accumulate there and can eventually cover the sparger rings. Cleaning requires sludge drawdown.

A. FREE GAS LIFT MIXING

B. FINE BUBBLE LANCES MIXING

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

TYPES OF GAS MIXING

FIGURE 15

C. SPARGER RING MIXING

D. DRAFT TUBE MIXING

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
TYPES OF GAS MIXING (cont.)

FIGURE 15

Placement of the sparger rings at or below mid-point of the digester has resulted in some success over the previous high maintenance.

4. Confined release of gas bubbles within a draft tube inside the digester. (Figure 15D)

Similar problems occur as with the sparger rings since little bottom velocity is created. The tube can be connected at the roof reducing maintainance problems associated with sludge drawdown.

The latter three systems can be designed to induce strong mixing to overcome surface scum accumulation. Normally, it will remain a routine problem. One device used at the Columbus, Wastewater Treatment Plant for scum control involved gas recirculation through pipes with outlets at various depths below the scum layer. (B.17)

Caution should be taken to avert excessive gas mixing which can create serious foaming problems and waste energy.

Studies by Lemke (I.4), Zoltak and Gram (E.4) and the Los Angeles County Sanitation District's Wastewater Treatment Plant (P.14) have shown that 0.3 Horsepower (Hp) is needed to mix 1000 cubic feet (cu. ft.) of digester capacity. Work in Orange County, California, showed that 0.20 Hp/1000 cu. ft. was needed. The Federal Environmental Protection Agency recommends 0.25 Hp/1000 cu. ft. (D.2) This power requirement is for the compressor motor and does not consider pipe friction losses or the compressor efficiency.

Studies on energy transfer of gas into the sludge have been done by Camp and Stein (P.4), Fair and Geyer (P.8), Buzzell and Sawyer (B.4), Vesilind (A.13), Dick and Ewing (C.3), Hatsfield (P.11) and Walker (B.26). Although the reference materials is in the bibliography, transfer efficiencies of gas will not be discussed in this report.

Approximately, 5-7 scfm (standard cubic feet per minute) per 1000 cu. ft. of digester gas at 6 psig (pounds per square inch, gage pressure) is sufficient for adequate gas mixing. (A.11) Work in Los Angeles County showed this amount of gas to vary with depth of gas injection.

One drawback common to all methods of gas mixing is its dependency on the production of digester gas. This can make gas mixing unreliable in smaller installations.

MECHANICAL MIXING. Mechanical mixing equipment can be located inside (internal) or outside (external) of the digesters. Internal mixing is accomplished with propellers or a combination draft tube/propeller. External mixing is accomplished with pump recirculation of the sludge.

Internal Mixing. Propeller mixers (Figure 16A) are normally roof mounted. Usually two or three units are suspended from the roof of a fixed cover digester with the shafts submerged 10 to 12 feet into the sludge. An electric motor drives the propeller which stirs the sludge. Maintenance problems associated with rag accumulation around the propeller are common.

Draft tube/propeller combinations can be in single (Figure 16B) or multiple units. This combination improves vertical mixing. The motor is usually reversible which helps but does not eliminate problems associated with rag accumulation around the propeller.

Two units running in opposite direction, either propeller or draft tube/propeller, help break up the scum layer. Otherwise, mixing occurs just below the scum blanket when the liquid level is kept constant.

A. PROPELLER MIXING

B. PROPELLER / DRAFT TUBE MIXING

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

INTERNAL MECHANICAL MIXING

FIGURE 16

Draft tube/propeller combinations can be side mounted. This method of installation helps keep grit off the tank floor but to remove or clean the equipment the digester has to be drawn down. Shaft seals tend to leak and are high maintenance items.

Other disadvantages associated with internal mixing include equipment abrasion from the sludge causing shaft bearing failure, and corrosion of equipment by the hydrogen sulfide in the sludge.

Large diameter, slow speed turbines (15-30 rpm) are more efficient mechanical mixers than the high speed propeller mixers. Lower operation and maintenance costs are also associated with the turbines.

External Mixing. External mixing (Figure 17) is actually pump recirculation mixing. Although this is more common in small installations, its use is increasing in larger facilities when combined with external heat exchangers. The heated sludge when mixed with the raw sludge can improve the heat blending in the digester. It helps avoid temperature shock to the microorganisms, thus, reducing the possibility of shortcircuiting. Pump recirculation can be advantageous in combination with other mixing systems.

At the Colorado Springs Wastewater Treatment Plant, centrifugal recirculation pumps (common in larger facilities) with directional nozzles inside the digester have successfully overcome surface scum problems and eliminated temperature stratification. (I.12)

In Dayton, Ohio, recirculation of sludge along the digester sidewall (peripheral mixing) added about 6 percent capacity to the overall sludge handling process. (K.2) Digester gas production was increased 13 percent. This may have been the result of more volatile

PUMP RECIRCULATION

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
EXTERNAL MECHANICAL MIXING

FIGURE 17

solids destruction due to the improved mixing efficiency. There was no evidence that the rate of gas production per pound of volatile solids destroyed had sustained any substantial increase.

There are maintenance problems associated with pump recirculation; namely, abrasion of the pump packing, shafts, wearing rings and impeller; and clogging.

Table 10 presents a comparison of the three major methods of digester sludge mixing. No one method can be selected as better since they're all dependent on sludge characteristics and whether additional solids handling processes are provided. Several studies using dyes and radioactive tracers have shown that, regardless of the method of mixing, the digester contents are never actually completely mixed; as might be accomplished within controlled laboratory conditions. Furthermore, the degree of mixing provided is directly related to the total power delivered to the contents of the digester tank, regardless of the mixing method. (I.13)

ANAEROBIC CONTACT PROCESS

An alternative method of mixing, called the anaerobic contact process, was studied in detail by Dietz, Clinebell and Strub in 1966. (B.8) The idea was first suggested by Schroepfer and Ziemke in 1959. (B.23) They proposed the recycling of concentrated sludge from the secondary digester, similar to returning activated sludge. Both the raw and recirculated sludge passed through a heat exchanger before returning to the digester. This process helped to maintain a high concentration of methane formers in the primary digester. Figure 18 is a schematic of this process as defined by Dietz, Clinebell and Strub. The benefits of

TABLE 10

COMPARISON OF THE VARIOUS
METHODS OF DIGESTER MIXING

	GAS MIXING	INTERNAL MECHANICAL MIXING	EXTERNAL MECHANICAL MIXING (Pump Recirculation)
ADVANTAGES	Low capital costs Low power cost	No scum problems	Low capital costs No mechanical equipment inside digester No scum problems
DISADVANTAGES	Scum problems Plugging High Operation and maintenance costs Possible corrosion	High capital costs High power costs High operation and maintenance costs Proppeller ragging	Plugging High power costs High operations and maintenance costs

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
ANAEROBIC CONTACT MIXING

FIGURE 18

recycling are only achieved when the returned sludge solids concentration is greater than that in the digester. Entrapped gases may cause dewatering problems. Consequently, vacuum degasifiers are sometimes provided. (B.8, B.23) Although used successfully for digesting industrial wastes, the anaerobic contact process has had little application to municipal wastewater because of its potential dewatering problems.

Data from a bench-scale study by Ghosh, Conrad and Klass (0.5) showed the anaerobic contact process to produce, on the average, 15.7 cubic feet of digester gas per pound of volatile solids destroyed with 65.9 percent methane content. These values show relatively no difference in gas production from other, more conventional mixing systems. Although the solids retention time (SRT) was reduced, it remains questionable whether full scale application would produce similar lower SRTs.

HEATING

Along with mixing, heating is one of the more important operational controls for optimizing digester gas production. Not only must heat be provided for the incoming sludge but the heat of the digester sludge must be maintained within plus or minus one degree Fahrenheit (plus or minus 1) of the optimum temperature mode (mesophilic or thermophilic). (A.11) Too high a temperature can pasteurize (partially sterilize) the bacteria. Too low a temperature can create short circuiting; regardless of the mixing efficiency. Consequently, the cold, dense raw solids sink to the bottom and are withdrawn only partially digested. (B.27) Heat losses through walls, floors, the digester cover and piping should be minimized if the net available gas is to be optimized.

The methods used to transfer heat into the sludge include the following.

1. Internal heating: heat exchangers placed inside the digester.

This is the least used method today. A long history of problems associated with cooking of sludge and caking the pipes have resulted in poor heat transfer and high maintenance.

2. Direct Flame: hot combustion gases are passed through the sludge.

This is the newest and, as yet, least known method. There is a potential explosion hazard from the highly combustible methane gas and pasteurization of the microorganisms from the high temperatures.

3. Direct Steam: steam injection directly into the sludge. Steam is produced in separate boilers or recovered from the vapor phase cooling of engines. Scaling can be a major maintenance problem. Steam condensation may produce higher moisture content in the digested sludge.

4. External Heating: external heat exchangers through which sludge is circulated.

This is the most common method with the least problems.

INTERNAL HEATING. Hot water circulation through horizontally mounted coils along the interior tank wall of the digester was the first method used to heat digester sludge. (P.22) In today's practice, pipe heating coils can be mounted horizontally or vertically. The water temperature has to be kept between 120 degrees Fahrenheit and 130 degrees Fahrenheit to maintain the sludge temperature within the mesophilic range. But corrosion problems can occur in the boiler. This

can be partially overcome by raising the water temperature to 180 degrees Fahrenheit and tempering the hot water with cool water returned from the digester. Corrosion can still occur in the boiler and in horizontally mounted pipe coils which are difficult to inspect and clean.

Vertically mounted coils, on the top of the digester tank, are less prone to scaling and can be removed for maintenance, if necessary. High boiler-water temperature is still required.

DIRECT FLAME HEATING. Direct flame heating of digester sludge is accomplished by burning gas (usually digester gas) in a specially designed combustion chamber and diffusing the hot gases into the sludge in separate, small heating tanks. Sludge is pumped into these tanks before digestion. The gas is at 2400 degrees Fahrenheit; but the transfer of heat to the sludge reduces the temperature to around 125 degrees Fahrenheit before it enters the digester. Overall thermal efficiency is about 85 to 90 percent. (P.22, P.21) Except for its application at a Wastewater Treatment Plant in Philadelphia in the 1950's, direct flame heating has rarely been used.

DIRECT STEAM HEATING. Steam injection has met with more favorable results than direct flame or internal heating. The main advantage is that it requires very little equipment. The steam is produced in separate boilers or can be recovered from the vapor phase cooling of engines. The steam is usually discharged through ports submerged about 12 feet into the digester. This helps to muffle the noise.

Direct steam injection can cause scaling. Besides, the digested sludge usually becomes diluted. This method requires 100 percent boiler make up water. Water softening may be needed, increasing the cost for this process considerably. Furthermore, if there are problems with digester mixing, stratification of the sludge can easily occur.

EXTERNAL HEATING. One of the oldest and probably most widely used method of heating sludge is the external heat exchanger. It has had successful use in England since 1923. In the United States it replaced the earlier use of internal heating coils. (B.27) The heat exchanger is popular because it combines the advantage of good scum control with the elimination of the vulnerable internal piping. The main disadvantage is that, if only one digester is used, sludge recirculation must be stopped to allow for liquid-solids separation and supernatant withdrawal.

There are three common types of external heat exchangers: water bath, jacketed pipe and spiral.

For the water bath heat exchanger, boiler tubes and sludge piping are located in a common water-filled container. Heat transfer is increased by using a pump to circulate the hot water across the sludge pipes. The heat exchanger and boiler are often located in a single unit which increases the possibility of an explosion.

For the jacketed pipe heat exchanger, hot water is pumped counter-current to the sludge flow through concentric water pipes surrounding the sludge pipe.

For the spiral heat exchanger, the hot water is also pumped counter-current to the sludge but the passageways for both water and sludge are cast in a spiral. (A.11)

In general, the heat transfer coefficients for the three types of external heat exchangers range from 150 to 275 Btu/hr/sq.ft./°F. Wesner claims that this Btu transfer rate substantially declines within a few months after plant start-up due to sludge caking and water impurities deposited along the pipe's interior wall. (I. 1)

Figure 19 shows a typical piping arrangement used to control the hot water supply for a jacketed pipe or spiral heat exchanger. The supply water temperature is kept below 155 degrees Fahrenheit to avoid sludge caking. This system allows the heat source to be remote from the heat exchanger. This, in turn, is a safer system and supports waste heat recovery.

Separate heat exchangers should be considered for each digester. An additional exchanger should be considered for the raw sludge whenever economically feasible. (A.11)

The following sections will address the various sources of heat, heat recovery and heat losses which affect overall gas production.

HEAT SOURCES

The hot water or steam used to heat the digester is often generated in a boiler. The boiler is usually fueled by digester gas with an alternative fuel source (natural gas, propane or fuel oil) used for back-up. Natural gas is most commonly used since it can be easily blended with the digester gas with minimal equipment adjustment. (A. 11) Frequently, the waste heat from digester gas-powered engines is sufficient to meet the digester heating requirements. The engines can be cooled using a forced draft system or a natural draft system. Forced draft cooling occurs when water is pumped through the engine. Natural draft or ebullient cooling occurs when water is vaporized and circulated

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
HEAT SOURCE SCHEMATIC
FOR SLUDGE HEAT EXCHANGER

FIGURE 19

without pumping. Ebullient cooling provides greater heat recovery (18 to 20 percent), and longer engine life. (D.13) A combination exhaust silencer and heat recovery unit can produce 10 to 13 percent additional heat. To prevent acid corrosion, recovered exhaust gasses should be kept above 400 degrees Fahrenheit.

Heat for the heat exchanger boiler can also be generated by solar panels (H.5) and heat recovery from pure oxygen aerobic digesters (B.9).

HEAT RECOVERY

The use of heat recovery equipment from the exhaust gases of digester gas-driven engines can increase the overall engine efficiency from 30 to 45 percent. In large installations, heat recovery readily provides enough energy to heat the digesters.

Typcially, 30 percent of recoverable energy, in the form of steam, is obtainable from the engine's jacket water which is used for cooling the engine. The process is called ebullient cooling. The recovered steam can be used as an energy source for a heat exchanger. An additional 15 percent of recoverable energy is possible from the exhaust stack of the engine. Water can be heated by the hot exhaust gas and returned to the cooling system for the engine. Consequently, a total of 45 percent additional energy is possible from heat recovery.

At the Hyperion Wastewater Treatment Plant in Los Angeles, heat recovery from their internal combustion (IC) engine, running on digester gas, adds over 30 percent to the overall efficiency of the engine. (L.1) Studies by Culp of serveral heat recovery systems show that an average of 25 percent additonal energy can be provided from heat

recovery of internal combustion engines. (A.12) This investigation covers plant capacities from 1 to 400 MGD. The Value Engineering report for the Las Vegas Wastewater Treatment Plant expansion shows that a 1700 horsepower engine driven by digester gas can provide, through heat recovery, enough energy to heat all six of the treatment plant's 66-foot diameter digesters. (I.5)

HEAT LOSSES

The amount of heat lost throughout the digestion system depends on the tank configuration, type of cover, construction materials, the air temperature and desired sludge temperature within the digester, the groundwater conditions, and insulation.

TANK CONFIGURATION. Anaerobic digestion tanks can be either cylindrical, rectangular or egg-shaped. (Figure 20). (A.11)

The most economical design, so far as heat loss is concerned, is the cylindrical shape when the diameter equals the depth. (J.4) The diameter of cylindrical tanks range from 20 to 125 feet, on the average, with a side water depth of 20 to 40 feet. The least expensive tank with diameter less than 65 feet is constructed of reinforced concrete. Internal reinforcement, like post tensioning rods, is more expensive. Steel has been used for tanks larger than 90 feet in diameter, but the cost is almost prohibitive.

Typically, the floor of a digester is conical. Recently, waffle-bottom floors have been designed to minimize grit accumulation. Although capital costs are higher, cleaning is reduced, saving in maintenance costs. Waffle-bottom digesters exist at the East Bay Municipal Utility District's Wastewater Treatment Plant in Oakland, California (B.8, B.3), and in other plants in Tacoma, Washington and Portland, Oregon.

A. RECTANGULAR

CONICAL BOTTOM TYPE

WAFFLE BOTTOM TYPE

B. CYLINDRICAL

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

DIGESTER TANK CONFIGURATION

FIGURE 20

C. EGG-SHAPED

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

DIGESTER TANK CONFIGURATION (cont.)

FIGURE 20

Rectangular shaped tanks provide a more simplified construction and are more efficient where space is limited. Mixing is usually a problem and dead spots often occur at the corners. They are not very common today.

Egg-shaped tanks have been used in Germany for over 20 years. They are less common in the U.S., first appearing in Kansas City, Missouri in the mid-1970's. They've recently been constructed at the Terminal Island Wastewater Treatment Plant in Los Angeles, California (P.21) The unique shape is aimed at eliminating grit accumulation on the bottom and minimizing scum accumulation on the top. Mixing is often accomplished with gas sparger rings and sludge recirculation. These tanks have the highest construction costs but remain unproven with regards to savings in operation and maintenance costs.

CONSTRUCTION MATERIALS. As mentioned previously, most digester tanks are constructed of concrete masonry; although, steel is used for larger installations. Digester covers are either reinforced concrete or steel. Sludge piping is usually of cast iron, sometimes lined with glass to alleviate grease problems, or ductile iron. Gas lines can be cast iron or wrought iron. Galvanized steel is sometimes used, also. Hot water lines are made of cast iron, wrought iron, galvanized steel or copper. Digester heating coils are usually cast iron, extra heavy wrought iron or copper.

The difference in construction materials affects the heat-loss rate, measured in Btu per hour. This heat loss is a function of the heat transfer coefficient which varies depending on the type of material. Heat loss coefficients for some typical materials are shown in Table 11. The coefficient, U, is in Btu per sq.ft. per hour per

TABLE 11
HEAT TRANSFER COEFFICIENTS

MATERIAL	THICKNESS	COEFFICIENT, U Btu/sq. ft./hr/°F
CONCRETE WALL (Insulated)	12 inches	.18 - .35*
CONCRETE WALL (No Insulation)	12 inches	0.86
CONCRETE ROOF	6 inches	0.50
FIXED STEEL COVER	1/4 inch	0.91
FLOATING COVER (Wood)	-	0.33
CONCRETE FLOOR (Exposed To Wet Earth)	12 inches	0.11
CONCRETE FLOOR (Exposed To Dry Earth)	12 inches	0.06

* depends on the type of insulation

degree Fahrenheit and depends on the thickness of the material. Comparison of these coefficients reveals that the heat loss through the floor of the digester is a relatively small part (0.06, depending on the groundwater table elevation) of the total potential heat loss from other areas of the digester. If the side walls are insulated, the heat loss can be minimized. About 65 percent of the total heat-loss occurs through the digester covers.

TYPES OF COVERS. Digester covers are either fixed, floating or gasholding.

Fixed covers (Figure 21) are made of steel, reinforced concrete or fiberglass reinforced polyester (FRP). They are usually dome shaped and are designed to maintain a constant water surface level. Concrete is most susceptible to cracking; especially in areas where climatic changes are drastic. A gas dome is often provided on flat or conical covers so that when the liquid volume changes, gas, and not air, will be drawn into the digester. (0.6)

Floating covers (Figure 22) are more expensive but provide flexibility in operating the digester. They reduce explosion hazards and can help control surface scum accumulation. They can be of the pontoon (Wiggins) type or the Downes type. (B.27). Both types float directly on the liquid, and are usually made of steel with built-in gas holding capacity instead of a separate gas dome. They have an effective travel distance of 2 to 8 feet. Downes covers contribute to better surface mixing. They reduce the area of the cover exposed to corrosive gases and are more structurally stable.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

FIXED DIGESTER COVER

FIGURE 21

WIGGINS TYPE

DOWNES TYPE

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
FLOATING DIGESTER COVERS

FIGURE 22

Floating gas holding covers (Figure 23) are also available. But because of the low storage pressure available, only about 3 to 6 hours of storage capacity can be obtained for average gas production rates. (A.11) Better storage is achieved by compressing the gas and providing a separate storage tank. These covers are the least stable since support is provided by a cushion of gas rather than liquid. On the other hand, the gas cushion does provide good insulation which reduces heat loss through the cover.

INSULATION

To minimize the heat loss in a digestion system, insulation of the digester tank walls, floor, cover and pipes should be considered. Insulating the tank walls and floor with earth is the most economical method of heat loss reduction; which is relatively small in these areas. Additional insulation is provided when an air space and masonry facing is constructed on the exposed tank wall or placing a manufactured insulating material around this portion of the wall. (J.4)

Insulating the cover is most efficient with a lightweight material. A one inch layer of polyurethane foam is often recommended. This material is inexpensive and structurally stable; as well as providing peripheral buoyancy of the cover. Cover stability is enhanced when the buoyancy is at the periphery rather than the center. Both cover and tank can be insulated with glass wool, corkboard, foam glass (polyurethane), lightweight concrete or air. Some facilities have used a PVC liner for cover insulation which seems to work as well as the polyurethane foam.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
GAS HOLDING DIGESTER COVER

FIGURE 23

Sludge piping and hot water transmission piping might also be insulated. Corrugated asbestos has been used in the past but because of its carcinogenic related problems it is not recommended. Mineral wool and magnesia are available but the lightweight plastic foams are inexpensive and provide about the same results.

CHAPTER III

PROCESS SELECTION

The variables affecting digester gas production were discussed in the previous chapter. An economical and technological comparison of each variable is presented below.

SLUDGE THICKENING

A comparison of the sludge thickening techniques is presented in Table 12. Sludge thickening benefits digester gas production by reducing the moisture content and increasing the concentration of volatile solids in the digester. However, volatile solids greater than nine percent can create ammonia toxicity in the digesters.

Adding sludge thickening to existing wastewater treatment plants is relatively easy. The energy requirement varies depending on the thickening method. Gravity thickeners require the least power; the imperforate bowl centrifuge requires the most. Nonetheless, only a small amount of the total plant energy requirement is attributed to thickening.

Using a digester for thickening has no proven benefits and is very costly compared to the other alternatives. For these reasons, it will not be considered further.

Elutriation, ultrafiltration and sludge lagoons are not commonly used for thickening. Consequently, little information is available to guarantee their performance and reliability. Capital costs for each are high. These systems will not be considered further.

The imperforate bowl centrifuge has fairly high capital and operating costs. Chemicals are required to obtain the high solids

TABLE 12
COMPARISON OF SLUDGE
THICKENING METHODS (A.12, C.8)

TYPE OF SLUDGE HANDLED	GRAVITY THICKENING	DISSOLVED AIR FLOTATION	CENTRIFUGES			ELUTRIATION	ULTRA- FILTRATION	SLUDGE LAGOON	DIGESTER
			SOLID BOWL	DISC NOZZLE	IMPERFORATE BOWL				
	PRIMARY AND COMBINED	BIO- LOGICAL	BIO- LOGICAL	INDUS- TRIAL	INDUS- TRIAL	BIOLOGICAL OR COMBINED	BIOLOGICAL	BIOLOGICAL OR COMBINED	PRIMARY AND/OR BIOLOGICAL
OPERABILITY	EASY	COMPLEX	DIF- FICULT	DIF- FICULT	DIF- FICULT	N.A.	N.A.	EASY	N.A.
MAXIMUM PERCENT THICKENING OF SLUDGE	3-5	4-5	7	5	9-10	N.A.	6	N.A.	N.A.
ODOR PROBLEMS	YES	NO	NO	NO	NO	N.A.	N.A.	YES	NO
CAPITAL COSTS	HIGH	MODERATE	LOW	MODERATE	HIGH	HIGH	HIGH	HIGH	HIGH
OPERATING COSTS	LOW	MODERATE	MODERATE	HIGH	MODERATE	LOW	HIGH	LOW	N.A.
ADD POLYMER TO INCREASE SOLIDS CAPTURE	NO	YES	YES	NO	YES	YES	NO	NO	NO
MAINTENANCE COSTS	LOW	MODERATE	HIGH	HIGH	MODERATE	LOW	LOW	LOW	N.A.
RELIABILITY	GOOD	GOOD	FAIR	POOR	POOR	N.A.	N.A.	POOR	UNPROVEN
FLEXIBILITY	GOOD	FAIR	POOR	POOR	FAIR	N.A.	N.A.	FAIR	GOOD

N.A. - INFORMATION WAS NOT AVAILABLE

capture and reliability is poor. It's application has been proven for industrial sludges only. Consequently, this method will not be considered further.

The disc nozzle centrifuge, although proven successful with industrial sludge thickening, has had only limited application to municipal sludges. Its flexibility and reliability in treating municipal sludge is poor. Operating costs can be high. It will not be considered further.

Gravity thickening, dissolved air flotation and solid bowl centrifuge can achieve good quality thickening of municipal sludges. All have proven technologies and are used today in wastewater plants. They will be considered further and a detailed economic comparison will be made later.

DIGESTER MIXING

A comparison of the various methods of digester mixing is presented in Table 13. Uniform and complete digester mixing provides optimum contact between the organic material (the food) and the microorganisms which, in turn, reduces the organic material. In this case the organic material is measured as volatile solids. As described in Chapter I, the breakdown of the organics yields methane and carbon dioxide. The greater the volatile solids reduction, the more methane is produced. Mixing was previously broken down into two categories; gas mixing and mechanical mixing.

Gas mixing is a viable alternative. Four methods of gas mixing were previously described. One of these methods, large bubble orifices, has exhibited a poor quality of mixing with moderately high

TABLE 13

COMPARISON OF DIGESTER
MIXING METHODS

	GAS MIXING			MECHANICAL MIXING			ANAEROBIC CONTACT PROCESS
	ORIFICE	LANCES	SPARGERS	INTERNAL		EXTERNAL	
				PROPELLER	DRAFT TUBE/ PROPELLER	PUMP RECIRCULATION	
QUALITY OF MIXING	POOR	GOOD	GOOD	POOR	GOOD	GOOD	GOOD
MAINTENANCE PROBLEMS	2,3	1,2	1,2,3	1,4	1,2,4	1	1
TANK DRAWDOWN FOR CLEANING	YES	NO	YES	POSSIBLE	YES	NO	NO
SPECIAL CONSIDERATIONS	REQUIRES COMPRESSOR	REQUIRES COMPRESSOR	REQUIRES COMPRESSOR	REQUIRES FIXED COVER	REQUIRES FIXED COVER	REQUIRES GOOD GRIT REMOVAL	REG. THICKENING MAY CAUSE DEWATERING PROB.
CAPITAL COSTS	LOW	LOW	LOW	HIGH	HIGH	MODERATE	HIGH
MAINTENANCE COSTS	MODERATE	LOW	MODERATE	MODERATE	MODERATE	HIGH	MODERATE
OPERATING COSTS	LOW	MODERATE	MODERATE	MODERATE	MODERATE	HIGH	HIGH
RELIABILITY	QUESTIONABLE	GOOD	GOOD	GOOD	GOOD	GOOD	QUESTIONABLE

1. Plugging or ragging
2. Foaming
3. Solids Accumulation
4. Wear in shaft bearings

maintenance costs. Because of its questionable reliability, it will not be considered further. Mixing with lances and sparger rings will be considered further.

Mechanical mixing, both internal and external, has also proven technically feasible. Propeller mixing is the least favorable because of its moderately poor mixing quality and high capital costs. It will not be considered further. The combination draft tube/propeller and pump recirculation provide good mixing quality with good reliability and will be considered further.

Although the anaerobic contact process has had limited application to wastewater sludges, it's potential for good mixing and heat distribution cannot be overlooked. Consequently, it will be considered further.

DIGESTER HEATING

A comparison of the various methods of digester heating is presented in Table 14. Whether the digester is operating within the mesophilic or thermophilic range, enough heat must be provided to maintain the maximum operating temperature to plus or minus 1 degree Fahrenheit throughout the tank. Heating can be directly applied to the sludge with steam or a flame. More common is the use of an external heat exchanger where sludge is circulated through a system of pipes and heat is transferred to it through adjacent counter-current flowing hot water.

Internal heat exchangers are seldom seen today except in old, smaller digesters. Because of the caking of sludge on the heating coils, heat transfer is low and maintenance is high. For these reasons, it will not be considered further. Direct flame is technically unproven as a reliable heating source. It will not be considered further.

TABLE 14
COMPARISON OF DIGESTER
HEATING METHODS

	INTERNAL HEAT EXCHANGER	EXTERNAL				
		DIRECT		HEAT EXCHANGER		
		FLAME	STEAM	WATER BATH	SPIRAL	JACKETED PIPE
HEAT TRANSFER	POOR	FAIR	GOOD	GOOD	GOOD	GOOD
WATER TEMP.	120-130°F	240°F (gas)	240°F	150°F	150°F	150°F
SCUM PROBLEMS	YES	YES	YES	NO	NO	NO
CAKING PROBLEMS	YES	NO	YES	NO	NO	NO
BOILER PROBLEMS	CORROSIONS	NO	100% MAKE UP WATER NEEDED	NO	NO	NO
OTHER HAZARDS	N.A.	EXPLOSIVE	STRATIFICA- TION IF POOR MIXING	EXPLOSIVE	NO	NO
CAPITAL COSTS	HIGH	MODERATE	MODERATE	HIGH	HIGH	HIGH
MAINTENANCE COSTS	HIGH	MODERATE	LOW	MODERATE	MODERATE	MODERATE
OPERATING COSTS	MODERATE	MODERATE	LOW	MODERATE	MODERATE	MODERATE
RELIABILITY	POOR	N.A.	GOOD	GOOD	GOOD	GOOD

Some success with direct steam heating exists. The hot steam can destroy microorganisms and moisture is added to the sludge defeating the purpose of previous thickening for volume reduction. It's use is very limited and does not warrant further consideration. Only the spiral and jacketed pipe external heat exchangers will be considered further. Both provide good heat transfer and are technologically reliable systems. Since the water bath method is being used less frequently, it will not be considered for further comparison.

DIGESTER CONSTRUCTION

Gas production may be affected by the digester tank configuration and the type of cover and materials used. Table 15 compares three digester tank configurations. Rectangular tanks are not recommended primarily because poor mixing is a common problem in the corners. Cylindrical tanks are the most widely used. Egg-shaped digesters have had very limited use in the United States. They are presently under construction at the Terminal Island WWTP in Los Angeles. There is no evidence that they produce more digester gas than cylindrical tanks and their capital cost is higher.

Digester covers are compared in Table 16. There is no comparative advantage between fixed and floating covers. Their application produces different results and should be considered on a case-by-case basis. Fixed covers should be used with roof-mounted mixing equipment or lances. Digester loading rates are not dependent on cover design but a floating cover or a gas holding cover provides more flexibility for gas storage. In cases where a combination of fixed and floating (or gasholding) covers are installed, a fixed cover is usually used for the primary digester.

TABLE 15
COMPARISON OF DIGESTER
TANK CONFIGURATIONS

	CYLINDRICAL	RECTANGULAR	EGG-SHAPED
MIXING PROBLEMS	NO	YES	N.A.
HEATING PROBLEMS	NO	NO	NO
CAPITAL COSTS	LOW	MODERATE	HIGH
MAINTENANCE COSTS	MODERATE	HIGH	LOW

TABLE 16
COMPARISON OF DIGESTER
TANK COVERS

	FIXED			FLOATING	
	STEEL	REINF. CONCRETE	FIBER GLASS	WIGGINS (STEEL)	DOWNES (STEEL)
HEAT LOSS*	HIGH	LOW	LOW	LOW	LOW
STRUCTURAL STABILITY	GOOD	FAIR	GOOD	FAIR	GOOD
CAPITAL COSTS	MODERATE	LOW	LOW	HIGH	HIGH
MAINTENANCE COSTS	MODERATE	MODERATE	LOW	MODERATE	LOW
FLEXIBILITY	FAIR	FAIR	FAIR	GOOD	GOOD

* Without insulation

SUMMARY

The following alternatives will be considered in the economic evaluation.

Sludge Thickening:

1. Mechanical Gravity;
2. Dissolved Air Flotation; and
3. Solid Bowl Centrifuge.

Digester Mixing:

1. Gas mixing using:
 - a. lances; and
 - b. sparger rings.
2. Mechanical mixing using:
 - a. draft tube/propeller; and
 - b. pump recirculation.

Digester Heating:

1. External Heat Exchanger:
 - a. spiral pipe; and
 - b. jacketed pipe.

Digester Construction (Including Insulation):

1. Cylindrical Tanks:
 - a. fixed covers; and
 - b. downes floating covers.

ECONOMIC ANALYSIS

INTRODUCTION. The purpose of the economic analysis is to compare the costs of each alternative within the above four groups. These alternatives have already been selected as providing the best technology for its particular process. The economic comparison should provide the final element for selecting the processes which provide the optimum digester gas production.

Determination of the costs for each process is partially based on the Construction Cost Index (CCI) as published in the Engineering

News Record (ENR). The index considers cost trends for construction materials and labor in 20 major cities across the country. Table 17 lists the index for each of these cities for March 31, 1981. Since the index changes frequently, this table is provided to show the proportion in costs between any two cities. The cost curves developed for the major equipment are based on an ENR Index of 4500, which is most representative of the City of San Francisco. These costs can be adjusted for any other city by dividing the CCI for that area by 4500. The second curve developed, when information was available, included labor required for the routine, normal maintenance of the equipment, (calculated in man-hours per year). The index for labor does not coincide with the Construction Cost Index. Labor requirements developed for the curves were based on a national average and do not distinguish between common or skilled labor. Table 18 lists the common labor index (CLI) and skilled labor index (SLI) for the 20 major cities.

The power consumption curve was based on continuous, 24 hour, operation of the equipment. The unit of energy used was either kilowatt hours or therms. Table 19 provides a list of phone numbers for the Public Utility Rate Departments responsible for developing the utility rate schedules in the 20 major cities. Rate schedules can be obtained free of charge by telephone request.

Some caution must be used when comparing the cost figures, particularly with the thickening processes. The type of sludge to be thickened is important. Primary sludge is effectively thickened with only the mechanical gravity thickener. Whereas, biological sludge can be effectively thickened with a dissolved air flotation thickener or a

TABLE 17
ENR CONSTRUCTION COST INDEX
MARCH 31, 1981

CITY	CCI
San Francisco, CA	4386
Los Angeles, CA	4084
Seattle, WA	3925
Cleveland, OH	3871
Detroit, MI	3796
New York City, NY	3775
Cincinnati, OH	3610
St. Louis, MO	3578
Kansas City, MO	3530
Pittsburgh, PA	3500
Chicago, IL	3473
Minneapolis, MN	3244
Philadelphia, PA	3234
Boston, MA	3163
Denver, CO	2947
Baltimore, MD	2896
New Orleans, LA	2773
Dallas, TX	2686
Atlanta, GA	2627
Birmingham, AL	2576

TABLE 18

ENR LABOR INDICES
MARCH 31, 1981

CITY	CLI	RANK	SLI	RANK
San Francisco	344	5	314	3
Los Angeles	356	1	336	1
Seattle	357	2	327	2
Cleveland	315	12	294	9
Detroit	313	13	314	4
New York City	257	20	250	20
Cincinnati	339	8	314	5
St. Louis	294	18	262	19
Kansas City	357	3	308	6
Pittsburgh	328	10	282	13
Chicago	298	16	289	10
Minneapolis	283	19	277	16
Philadelphia	310	15	267	17
Boston	295	17	265	18
Denver	311	14	281	14
Baltimore	342	1	281	15
New Orleans	348	4	297	8
Dallas	344	6	302	7
Atlanta	330	9	282	12
Birmingham	317	11	289	11

TABLE 19

PUBLIC UTILITIES
RATE DEPARTMENT

CITY	GAS	ELECTRIC
San Francisco	415-781-4211	SAME
Los Angeles	213-620-2564	SAME
Seattle	206-625-3000	--
Cleveland	216-664-3340	SAME
Detroit	313-965-8000	237-8000
New York City	212-460-4600	SAME
Cincinnati	513- ---	--
St. Louis	314-621-6960	342-1000
Kansas City	816-221-3300	471-7000
Pittsburgh	412-281-6660	456-4000
Chicago	312-431-4000	294-4321
Minneapolis	612-372-4664	330-5500
Philadelphia	215-796-1000	841-4000
Boston	617-742-8400	424-2000
Denver	303-623-1234	SAME
Baltimore	301-234-5000	SAME
New Orleans	504-586-2121	SAME
Dallas	214-741-3750	653-1311
Atlanta	404-522-1150	325-4001
Birmingham	205- ---	250-1000

centrifuge. But, the design basis for these two units is not comparable. A DAF thickener is designed based on surface area; whereas, a solid bowl centrifuge bases its design on capacity (a function of the centrifuge speed). Costs for manpower and power consumption are similarly based.

On the otherhand, the costs for digester mixing - gas mixing, mechanical mixing, pump recirculation -- can be directly compared on the figures since each is directly related to digester surface area.

The capital cost for each process is based on the manufacturers price for the process equipment. Additional costs, such as installation, engineer's fee, legal fees, and the like, were not included. The costs are often a percentage add-on to the equipment cost and can vary. For comparison purposes, the costs provided should be legitimate.

SLUDGE THICKENING

Mechanical Gravity Thickening is considered the best method for thickening primary or combined (primary and biological) sludge from those methods compared in Table 10. Costs are provided in Figures 24, 25 and 26. Figure 24 (P. 19) shows the construction costs (for the process equipment) for mechanical gravity thickeners. The costs for generating curve A in Figure 24 were developed by direct manufacturer contact and from several project costs in California. These costs were taken to the common ENR base index of 4500. Curve C was developed from an EPA study of construction bid data for wastewater treatment plants throughout the United States. (P. 19) The equation, $\text{Cost (C)} = 3.28 \times 10^4 \times \text{Flow (Q)}^{1.10}$ was developed from this study. Curve B uses this

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR
MECHANICAL GRAVITY THICKENERS**

FIGURE 24

equation but is adjusted to an ENR index of 4500. Actual construction costs for a mechanical gravity thickener is expected to fall between Curves A and B for an ENR index of 4500. Figures 25 (A.11) and 26 (A.11) provide the manpower (hours per year) and energy (KWH per year) required to maintain and operate a mechanical gravity thickener. Surface area (square feet) is used for sizing the thickener since, in most design cases, the loading rate, either mass loading (pounds per square foot per hour) or hydraulic loading (gallons per square foot per hour) will be the determining factor for sizing the unit.

Dissolved air flotation thickening is considered the best method for thickening separate biological sludges. Costs are provided in Figures 27, 28 and 29. Curve A, Figure 27 was based on actual projects constructed within the last 10 years in California and adjusted to the ENR index of 4500. Curve C was based on the EPA study of construction bid data for wastewater treatment plants throughout the United States. (P. 19) The equation, $\text{Cost (C)} = 2.99 \times 10^4 \times \text{Flow (Q)}^{1.14}$ was developed from this study. Curve B, using this equation, was adjusted to an ENR index of 4500. Actual cost is expected to fall somewhere between Curves A and B. Figures 28 (A.11) and 29 (A.11) provide an estimate of the manpower (hours per year) and power consumption (KWH per year) required to operate and maintain a dissolved air flotation thickener.

Centrifugal thickening of waste activated sludge has met with some success but is not as reliable as mechanical gravity thickening or dissolved air flotation. Of the three types of centrifuges reviewed, only the solid bowl centrifuge had sufficient thickening cost data available from its application to wastewater treatment. Because of the

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

**MANPOWER FOR
MECHANICAL GRAVITY THICKENERS**

FIGURE 25

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
POWER CONSUMPTION FOR
MECHANICAL GRAVITY THICKENERS

FIGURE 26

FEASIBILITY STUDY
FOR ALTERNATIVE FUEL PRODUCTION
**COSTS FOR DISSOLVED
AIR FLOTATION THICKENERS**

FIGURE 27

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**MANPOWER FOR DISSOLVED
AIR FLOTATION THICKENERS**

FIGURE 28

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**POWER CONSUMPTION FOR
DISSOLVED AIR FLOTATION THICKENERS**

FIGURE 29

limitation in particle size that can be thickened with a centrifuge, primary sludge is not considered. However, the future of centrifugal thickening is bright. Considering the small size of the equipment, it is most desirable at wastewater treatment plants where limited space is available. Construction costs for a solid bowl centrifuge are provided in Figure 30. (A.11) Costs are based on the unit's volume handling capacity in gallons per minute. Although loading rates are a key parameter for design, the lack of testing data has caused uncertainty in the optimum loading rate value. The maintenance cost in annual manpower hours is also based on handling capacity as shown in Figure 31. Power consumption, Figure 32, is estimated at 0.28 to 0.37 horsepower (.21 to .28 kilowatts) per gallon per minute flow rate. (A.11)

DIGESTER MIXING

The advantages and disadvantages of mixing a digester with gas or by mechanical methods have been discussed previously. The economic evaluation that follows is not compared to select the best alternative since both methods are relatively equal.

Gas Mixing. Gas mixing with lances has a slight edge over a sparger ring. A telephone survey of several wastewater treatment plants showed that the overall advantage of using lances was less maintenance than the sparger ring. The construction costs for a digester with gas mixing is shown in Figure 33. There is no significant difference in construction cost whether lances or a sparger ring is used for gas mixing. Power consumption is directly related to the motor horsepower used to drive the gas compressor. Sufficient mixing can be achieved with 0.2 to 0.3 horsepower per 1000 cubic feet of digester capacity

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

**COSTS FOR
SOLID BOWL CENTRIFUGE**

FIGURE 30

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

MANPOWER FOR SOLID BOWL CENTRIFUGE

FIGURE 31

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**POWER CONSUMPTION FOR
SOLID BOWL CENTRIFUGE**

FIGURE 32

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR ANAEROBIC DIGESTERS
WITH GAS MIXING**

FIGURE 33

(0.2 - 0.3 Hp/1000 ft.³). Our telephone survey also showed that the range of motor horsepower per 1000 cubic feet of digester capacity for lances was generally less (.05 to .16) than that for sparger rings (.10 to .30). The digesters surveyed ranged in size from 65 to 125 feet in diameter.

Because there are losses within the mixing system due to transfer and compressor efficiencies, and friction, the overall efficiency of gas mixing is about 50 percent. Operating costs for a gas mixed digester are provided in Figure 34. (A.1) These costs are based on a continuous mixing operation with gas release 20 feet into the digester. Motor efficiencies varied between 85 to 93 percent, depending on the motor size.

Mechanical Mixing. There are some minor differences between the draft tube/ propeller and pump recirculation methods of mechanical mixing. As indicated in Table 13, the internal draft tube/propeller costs less to operate and maintain; whereas, pump recirculation is less complicated, therefore, easier to operate. Additionally, it has a lower capital cost for smaller digesters, less than about 100 feet in diameter. Both provide about the same quality of mixing.

The external draft tube method can be constructed with a hot water jacket around the pipe which heats the sludge prior to its entering the digester. Similarly, pump recirculation can be constructed with an external heat exchanger installed just prior to the raw sludge entering the digester. Both methods are now operating successfully at several wastewater treatment plants across the country. The process was defined earlier in this Chapter as the Anaerobic Contact Process.

Construction costs for internal mechanical mixing are provided in Figure 35. Construction costs for pump recirculation are provided in

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**POWER CONSUMPTION FOR
DIGESTERS WITH GAS MIXING**

FIGURE 34

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR ANAEROBIC DIGESTERS
WITH MECHANICAL MIXING**

FIGURE 35

Figure 36. Comparing the two graphs shows that digesters larger than about 100 feet in diameter are less costly with mechanical mixing than with pump recirculation.

Operating costs for a mechanically mixed digester are given in Figure 37. Costs are, obviously, directly related to the size of the motor. Operating costs for pump recirculation mixing (basically sludge pumping) are given in Figure 38.

DIGESTER SLUDGE HEATING

Only the external heat exchanger was considered as the most viable method to heat sludge prior to entering the digester. The most unfavorable aspect of using a water bath exchanger is that the boiler tubes and sludge piping are located in a common water-filled container. Since the heat exchanger and boiler are combined in a single unit, the potential hazard of an explosion is increased over the other types of external heat exchangers. A description of the jacketed pipe heat exchanger and the spiral exchanger was provided earlier in this report. Both have similar heat transfer coefficients and the costs for operation and maintenance are indicated in Table 14.

Costs for such an external a heat exchanger include a boiler which is often fueled by digester gas but capable of using an alternative fuel (such as natural gas) to maintain continuity during periods of low digester gas production.

In wastewater treatment plants where digester gas is used to drive engines or generate electricity, the waste heat off this equipment can be recovered and used to heat the digester boiler. Typically, 18 to 20 percent of the heat can be recovered from an engine cooling system.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR ANAEROBIC DIGESTERS WITH
PUMP RECIRCULATION MIXING**

FIGURE 36

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**POWER CONSUMPTION FOR DIGESTERS
WITH MECHANICAL MIXING**

FIGURE 37

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**POWER CONSUMPTION FOR DIGESTERS
WITH PUMP RECIRCULATION MIXING**

FIGURE 38

A combination exhaust silencer and heat-recovery unit can be used to extract from an engine's exhaust an additional 10 to 13 percent of the heat. Making the total potential for heat recovery from 18 to 33 percent. (D.13)

The amount of heat required to raise the temperature of the raw sludge so it doesn't create a temperature shock to the already heated sludge in the digester varies with the total solids concentration of that raw sludge. This is shown in Figure 39.

Heat losses within the digester system will be discussed later along with digester insulation.

Costs for an external heat exchanger, including the boiler, are given in Figure 40. There is no significant difference between the cost of a spiral exchanger or jacketed pipe exchanger. There is a straight line cost relationship, Figure 41, for the boiler based on its capacity in BTUs per hour.

DIGESTER INSULATION

Insulation of a digester or piping does not, in itself, affect the production of digester gas. But it does have an indirect effect on the net digester gas available as an energy source because of an increase or decrease in heat losses.

Figure 42 shows the potential savings based on energy saved from reduced heat loss by insulating the digester. The energy saved in KWH is based on an ENR of 4500 and \$.051 per KWH. The net benefit is extended for a 20 year life cycle of the digester.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**EFFECT OF SOLIDS CONCENTRATION
ON THE RAW SLUDGE HEATING REQUIREMENT**

FIGURE 39

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR HEAT EXCHANGER
AND BOILER**

FIGURE 40

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

COSTS FOR BOILERS

FIGURE 41

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**ACCUMULATIVE SAVINGS FROM
DIGESTER TANK INSULATION**

FIGURE 42

CHAPTER IV

OPTIMIZING DIGESTER GAS PRODUCTION AT EBMUD WATER POLLUTION CONTROL PLANT

INTRODUCTION

The East Bay Municipal Utility District (EBMUD) Water Pollution Control Plant is located in Oakland, California. It is designed to treat 120 million gallons per day (MGD) of municipal and industrial wastewater. The plant location and service area are shown in Figure 43. This plant was chosen as an example to apply those processes previously selected as the most technically feasible and cost-effective for optimizing digester gas production. According to the flow schematic in Figure 2, the solids handling processes at EBMUD include: waste activated sludge thickening, anaerobic digestion, and sludge dewatering. A detailed process description and solids handling schematic are provided in Table 20 and Figure 44, respectively.

EXISTING PERFORMANCE

The primary sludge is not thickened. The waste activated sludge (WAS) is thickened with two solid bowl centrifuges. One is a low-speed centrifuge the other is a medium-speed centrifuge. Although the medium speed centrifuge uses twice as much energy (horsepower) to operate as the low speed centrifuge, it uses only one-third of the amount of chemicals. According to plant personnel chemical costs are very high. Consequently, the medium speed centrifuge is slightly favored in having less total operating and maintenance costs.

TABLE 20

EBMUD SOLIDS HANDLING PROCESSES

Process		Description
I. WASTE ACTIVATED SLUDGE (WAS) THICKENING (2 units)		
A. Low Speed Centrifuge		
1. Type		horizontal solid bowl
2. Size		200 hp - main drive 30 hp - backdrive
3. Capacity		400 gpm (design) at 4% cake solids and 85% capture
4. Chemical Addition		cationic polymer
B. Medium Speed Centrifuge		
1. Type		horizontal solid bowl
2. Size		400 hp - main drive 40 hp - backdrive
3. Capacity		400 gpm (design) at 7% cake solids and 90% capture
4. Chemical Addition		cationic polymer
II. DIGESTION		
A. Primary		
1. Type		heated, gas-mixed, with floating covers (high- rate)
2. Number		10
3. Size		diameter, 95 feet depth, 30 feet
4. Capacity		2 million gallons

TABLE 20 (Continued)

EBMUD SOLIDS HANDLING PROCESSES

Process	Description
5. Solids Loading	0.14 pounds of volatile solids per cubic foot per day
6. Hydraulic Retention Time	8 days
B. Secondary	
1. Type	unheated, unmixed, with floating covers (standard rate)
2. Number	2
3. Size	diameter, 95 feet depth, 30 feet
4. Capacity	2 million gallons

WASTE
ACTIVATED
SLUDGE

PRIMARY
SLUDGE

LOW SPEED
CENTRIFUGE

CENTRATE

SLUDGE

MEDIUM SPEED
CENTRIFUGE

CENTRATE

SLUDGE

SPIRAL
HEAT
EXCHANGER

HOT
WATER

BOILER

DIGESTER GAS

DIGESTED
SLUDGE TO
DEWATERING

DIGESTER

8 WAFFLE BOTTOM DIGESTERS
4 CONICAL BOTTOM DIGESTERS

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
PROCESSES AFFECTING
DIGESTER GAS PRODUCTION

FIGURE 44

IV-5

The primary sludge contains an average of 6.5 percent total solids. The corresponding volatile solids content averages 74 percent. The Waste Activated Sludge (WAS) contains 1.9 percent total solids. The corresponding volatile solids content averages 82 percent. The WAS is thickened to about 3 to 5 percent solids by centrifuging. (The volatile solids content does not change). The primary and WAS are then combined to produce the "feed sludge" to the digesters. This sludge contains about 4 to 6 percent solids with an average of 80 percent volatile solids. This 80 percent is a weighted average which takes into account the difference in volumes of primary and waste activated sludge. The ratio is approximately 1 to 3, primary to WAS.

There are twelve digesters. Presently, ten are used as primary digesters and two as secondary. These two secondary digesters are unheated and unmixed. The average amount of volatile solids reduced in the primary digesters for the period July, 1980 to June, 1981 is set forth in Table 21. The average daily volatile solids reduction was 56 percent.

Heat for the ten digesters is provided by four digester gas-fired boilers, each with a heating capacity of 10 million British thermal units (Btu) per hour. Each of the heated digesters has a spiral plate heat exchanger. Gas sparger rings are used for digester mixing.

Digester temperature is not recorded on the monthly data worksheets but plant personnel said that 95 degrees Fahrenheit was usually maintained.

The solids retention time (SRT) is not measured. Plant personnel use the hydraulic retention time (HRT) instead. The average HRT for the period from July, 1980 to June, 1981 was 35.6 days.

TABLE 21

VOLATILE SOLIDS REDUCTION

	Primary Sludge			Waste Activated Sludge			Combined	Digested	Reduction ₃ in VS lbs x 10 ³ %	
	TS in	VS in TS	VS	TS in	VS	VS	Sludge	Sludge		
	Sludge			Sludge			VS Portion	VS Portion		
	%	%	lbs x 10 ³	%	%	lbs x 10 ³	%	%		%
<u>1980</u>										
Jul	5.5	74.5	88.66	2.2	77.3	124.45	76.5	62.5	104	48.8
Aug	6.6	78.8	96.92	2.0	80.0	119.20	79.7	60.9	130	60.3
Sep	6.3	76.2	110.64	1.9	78.9	102.57	78.1	66.7	93	43.8
Oct	6.8	77.9	97.38	1.8	88.9	122.68	86.2	66.7	149	67.9
Ncv	6.5	76.9	83.82	1.9	84.2	128.83	82.3	65.0	128	60.1
Dec	6.6	62.1	73.90	1.8	83.3	116.62	77.6	63.6	94	49.6
<u>1981</u>										
Jan	6.8	75.0	81.00	2.0	80.0	136.00	78.9	63.6	116	53.3
Feb	6.5	76.9	57.68	1.9	84.2	107.78	81.3	60.0	108	65.5
Mar	6.3	73.0	57.67	1.6	81.3	113.82	77.8	61.5	93	54.4
Apr	6.5	73.8	42.80	1.8	83.3	163.27	81.1	64.0	121	58.6
May	6.4	71.9	49.61	2.4	79.2	190.08	77.4	61.5	128	53.4
Jun	<u>5.4</u>	<u>--</u>	<u>--</u>	<u>2.3</u>	<u>78.3</u>	<u>204.36</u>	<u>--</u>	<u>(60.0)</u>	<u>--</u>	<u>--</u>
<u>Avg</u>	6.44	74.27	76.37	1.94	81.9	135.81	79.7	63.3	115	56.0

TS - total solids
VS - volatile solids

OPTIMIZING GAS PRODUCTION

The processes selected in Chapter III as the best alternatives for optimizing digester gas production will now be considered for the EBMUD water pollution control plant.

SLUDGE THICKENING. The method of sludge thickening is restricted because of a lack of space at the site. Consequently, centrifugal thickening was selected. Gravity thickening of primary sludge was not considered. Thickening primary sludge would not significantly increase the amount of volatile solids in the digesters. Of the centrifugal methods available for thickening biological sludges, only the solid bowl method was recommended. It is also the method used at EBMUD. Low speeds are used to keep the solids level below 8 percent which eliminates the concern for ammonia toxicity.

Recommendation. The present method of thickening uses the best practical process and contributes to the optimization of digester gas production.

DIGESTER MIXING. Digester gas mixing is accomplished with sparger rings. The most serious problem encountered is foaming due to the large volume of waste activated sludge being fed to the digesters. This has been controlled somewhat by reducing the gas stream levels to the spargers. The waffle-bottomed digesters have helped in keeping down grit accumulation. Consequently, tank drawdown for cleaning has not been required.

Recommendation. No process changes in digester mixing would improve the potential for producing more digester gas. Although some facilities have experimented with raising the sparger rings to cut down the foam build-up, reducing the gas flow to the spargers has resulted in

considerably lower levels of foam. Also, mixing with pumps has been succesful in reducing foaming.

DIGESTER HEATING. Spiral plate heat exchangers are provided for each primary digester. The digesters are operated within the mesophilic range at a temperature of 95 degrees Fahrenheit. There is a direct relationship between temperature, solids retention time, and digester gas production (Figures 13 and 14, Chapte II). At 95 degrees Fahrenheit (35 degrees Centigrade), maximum gas production occurs within about 30 days. Because gas production is not measured at the EBMUD facility, it is difficult to determine whether the SRT, temperature, and gas production are optimum.

Most data generating studies relating digester sludge detention time and gas production use SRT. The EBMUD plant only measures HRT (hydraulic retention time) and there is no accurate correlation between the two.

The 56 percent reduction in volatile solids (Table 21) indicates better than average performance of the digesters. But since gas production is not metered, it is difficult to assess the efficiency of converting this amount of volatile solids reduction to digester gas. Table 22 shows the average daily amount of gas which could be available for three possible rates of gas production. From an earlier EBMUD study June 1981, 1.29 million cubic feet per day of digester gas was estimated to be EBMUD's current gas production. Interpolating with the production rates from Table 22, this would yield a rate of about 11 cubic feet per pound of volatile solids destroyed. This is lower than what might be expected with the high volatile solids reduction. The low yield may be

TABLE 22
DIGESTER GAS PRODUCTION

	<div>Lbs of VS Reduced x 10³</div>	Gas Production x 10 ⁶ ft ³ per day @		
		10 ft ³ /lbs VS red.	12 ft ³ /lbs VS red.	15 ft ³ /lbs. VS red.
<u>1980</u>				
Jul	104	1.04	1.25	1.56
Aug	130	1.30	1.56	1.95
Sep	93	0.93	1.12	1.40
Oct	149	1.49	1.79	2.24
Nov	128	1.28	1.54	1.92
Dec	94	0.94	1.13	1.41
<u>1981</u>				
Jan	116	1.16	1.39	1.74
Feb	108	1.08	1.30	1.62
Mar	93	0.93	1.12	1.40
Apr	121	1.21	1.45	1.82
May	128	1.28	1.54	1.92
Jun	--	--	--	--
<u>Avg.</u>	115	1.15	1.38	1.73

from digester sludge heat loss from the digester covers. This low temperature sludge would produce a low level of methane gas as well.

Recommendation. The SRT should be determined so an accurate correlation between HRT and SRT can be made. Also, it may be possible that the same volatile solids reduction will occur at a lower detention time (about 30 days). The potential benefit of reducing the detention time is that less thickening would be required. To accomplish this, the hours of operating the centrifuges could be reduced thus saving energy or the amount of chemical used could be reduced.

Experiment with maintaining the digester temperature at different temperatures around 95 degrees Fahrenheit to see if significantly more volatile solids can be reduced.

DIGESTER CONSTRUCTION. All the digesters are cylindrical in shape with floating covers. Since there is no roof-mounted mixing equipment, the floating covers present no operating or maintenance problems. The covers are of the Downes Type shown in Figure 22 in Chapter II. The air space between the top and bottom plates of the cover acts as an insulator. Gas accumulates in the center well where it is also drawn off.

As indicated in Table 11, Chapter II, different materials have different heat transfer coefficients. In determining the heat loss on the EBMUD digester covers, consideration has to be given to the two cover plates (steel bottom and wooden top) and the heat retained in the air space.

Recommendation. Determine whether it is cost effective to insulate the digester covers.

COST ANALYSIS FOR EBMUD MODIFICATIONS

Below is a summary of the previous recommendations made regarding optimizing digester gas production at the EBMUD Water Pollution Control Plant.

1. Continue waste activated sludge thickening with the solid bowl centrifuges. Do not thicken primary sludge.
2. Continue gas mixing of the digesters with the bottom mounted sparger rings.
3. Continue heating the raw sludge with the spiral plate heat exchangers.
4. Determine a correlation between solids retention time and hydraulic retention time. Reduce detention time by increasing water content of the raw sludge. This can be accomplished by reducing thickening.
5. Experiment with increasing temperature of the digester sludge to determine if volatile solids reduction increases at a different temperature.
6. Determine cost-effectiveness for insulating the ten primary digester covers.

COSTS. Each of the above six items will be evaluated for costs and payback based on the potential increase in digester gas production or energy saved.

1. Thickening. No modifications are recommended.
2. Mixing. No modifications are recommended.
3. Solids Retention Time (SRT). The modifications recommended require very little, if any, additional operating costs; and should be considered a part of the routine operations control. There is evidence that the high detention time (35.6 days) is caused by a lack of downstream solids handling capacity (i.e. dewatering and final sludge disposal). If this is remedied, then it is recommended to take one of the centrifuges out of service and use it for back-up. Plant data indicates that about 55 to 56 percent volatile solids reduction from 1980 and 1981 can be achieved within 32 days.

4. Temperature. The modifications recommended are part of routine operational control.
5. Heating. No modifications are recommended.
6. Insulation. The costs for digester cover insulation are evaluated in the following section.

COST EVALUATION OF DIGESTER INSULATION

STEP 1. DETERMINE HEAT LOSS FROM COVER. Cover loss from the ten primary digesters was calculated to be 171,900 Btu/hr/digester x 10 digesters x 24 hours/day:

41.256 million BTU per day

STEP 2. DETERMINE DOLLAR VALUE OF HEAT LOSS. The cost for purchasing a therm of natural gas varies somewhat within the Bay area. An overall average of \$0.40 per therm was determined from billing records for 1980 and is used for the following calculations.

$$\frac{41.256 \text{ Million BTU/day} \times 365 \text{ days/yr} \times \$0.40/\text{therm}}{100,000 \text{ BTU/therm}}$$
\$60,000/year

STEP 3. DETERMINE COST FOR DIGESTER COVER INSULATION. The cost for a semi-rigid, 1-inch thick, polyurethane (plastic) foam is about \$2 per square foot. The cover would have to be recoated every eight years at a cost of \$0.65 per square foot. The life of the material is 40 years. Cost of the insulation would be:

$$\begin{aligned} &\$2/\text{sq. foot} \times 7100 \text{ sq. feet/digester} \times 10 \text{ digesters} = \$142,000; \text{ plus} \\ &\$0.65/\text{sq. foot} \times 7100 \text{ sq. feet/digester} \times 10 \text{ digesters} \times 5 = \$230,750 \end{aligned}$$

\$373,000

Added to this cost would be installation, engineering fee, and legal and administrative costs which would bring the total capital cost to:

\$596,000

and the payback for this system would be ten years.

Even though this assumes that the price of natural gas will remain constant, it is a payback longer than what is considered reasonable. Consequently, insulating the digester covers is not recommended.

CHAPTER V

DIGESTER GAS SCRUBBING

INTRODUCTION

Digester gas, as shown in Table 3, Chapter I, has a heat value of between 590 and 650 British Thermal Units (Btu) per cubic foot of gas in the raw state. Consequently, with this BTU value, this gas is a potential energy source and has many applications. It can be burned directly to provide building heat, or burned in a boiler. The hot water produced can be used throughout the plant and, in particular, used in a heat exchanger to heat sludge for digestion. This is the most common use of digester gas in many wastewater treatment facilities, especially those treating less than one million gallons of wastewater per day (1 MGD).

In larger facilities, as the quantity of gas increases so does its uses. It can be used to power engines which can drive pumps and blowers or it can be used to generate electricity. It can also be sold to the local public utility or private industry. In order to do this in many cases, the gas must be cleaned to what is termed "pipeline quality". This means that it must contain no more than 10 parts per million (ppm) hydrogen sulfide (H_2S) and no more than 2 percent carbon dioxide (CO_2). Typical digester gas contains approximately 65 percent methane, 30 percent carbon dioxide and the 5 percent balance consists of hydrogen sulfide and other impurities. The process of removing the carbon dioxide and hydrogen sulfide is called gas scrubbing or gas cleaning. Since the scope of this report does not include cleaning digester gas to pipeline quality, only those processes which remove hydrogen sulfide or carbon dioxide or both are considered.

The extent to which digester gas is cleaned and a process selected is primarily based on two things: the amount (concentration) of contaminants in the gas and its use.

There are two reasons for removing hydrogen sulfide. One is for air quality control. Some areas of the country have restrictions on the amount of sulfur discharged to the atmosphere. The second is when hydrogen sulfide is removed from digester gas, the gas can be used to operate standard equipment. Otherwise, the equipment has to be specially manufactured to protect the internal parts from corrosion caused by the mixture of the hydrogen sulfide and moisture. Presently, many engines using digester gas do not need gas scrubbing since the hydrogen sulfide level in the gas is not very high.

Additionally, carbon dioxide is removed to increase the energy (BTU) value of the gas and to reduce the volume. As previously noted, about 30 percent of digester gas consists of carbon dioxide. The removal of the carbon dioxide becomes important when gas storage is required. The reduction in volume results in a cost savings for the storage equipment.

Much of the information on gas scrubbing processes was obtained from the gas industry. Their experience in landfill methane utilization technology is further developed than digester gas scrubbing in the wastewater industry. For a copy of a draft workbook on gas cleaning technology in the landfill industry, contact Mr. Russell C. Eberhart at Johns Hopkins University, (301) 953-7100, extension #3720.

GAS SCRUBBING TERMINOLOGY

The following definitions are intended to familiarize the reader with common terms used in the gas scrubbing techniques that will be evaluated later.

Absorption is the process whereby the gases, liquids or dissolved substances penetrate the surface of a liquid. The impurities from the gas stream are removed by dissolving or absorbing them in the liquid (absorbent). This can be accomplished by breaking the liquid up into tiny drops dispersed through a volume of gas (spray towers), or breaking the gas up into little bubbles passing through a volume of liquid (trayed towers), or breaking the liquid up into a multiplicity of slow-flowing films which form and re-form through a volume of gas (packed towers).

The absorption process usually generates heat. Physical absorption is more efficient than chemical absorption at higher pressures. It requires less heat than chemical absorption but the solvents are usually more corrosive. Solvent regeneration is usually less expensive with physical absorption processes.

Adsorption is a process whereby the gases, liquids or dissolved substances condense on the surface of a solid (or adheres to the surface). As a rule, adsorption occurs when the adsorbent is a solid. Physical adsorption occurs from weak forces of attraction between molecules. Consequently, it is easily reversible with a decrease in pressure. The adsorbed material is desorbed to the same extent that it was originally adsorbed. Chemical adsorption occurs from much stronger molecular forces which often results in the formation of chemical compounds. It is seldom reversible and, generally, requires very high temperatures to remove the adsorbed material.

Oxidation is the process whereby the molecules in a gas, liquid or dissolved substance loses electrons. Typically this occurs with the

addition of oxygen. Ozonation, another gas scrubbing process, is actually a form of oxidation whereby the gas, liquid or dissolved substance is ozonized. Ozone is a triatomic form of oxygen (O_3) that is normally a bluish irritating gas with a characteristic pungent odor.

GAS SCRUBBING PROCESSES

Table 23 lists the gas scrubbing processes that were reviewed for this report. A brief description of each process follows. Not all of the processes are applicable to the wastewater treatment industry. Therefore, a selection matrix has been provided at the end of this section to select those processes which will be further evaluated.

PHYSICAL ABSORPTION PROCESSES

SELEXOL PROCESS. The absorbing agent in this process is dimethyl ether of polyethylene glycol. This proprietary chemical, trade named Selexol Solvent, is noncorrosive and nontoxic, and has a low viscosity and a low vapor pressure. The proprietary process was developed and is licensed by the Allied Chemical Corporation. The Selexol Solvent is expensive (over \$9 per gallon) and its preferential absorption of propane-plus fraction makes the process mainly suitable to lean gas streams. The Selexol process is capable of scrubbing lean gases to pipeline specifications. Although the solvent can be regenerated, cost for the process is relatively high; by as much as 50 percent over other processes.

ESTASOLVAN PROCESS. The absorbing agent in this process is tributylphosphate. The process can remove both hydrogen sulfide and carbon dioxide. The flow scheme of the Estasolván process is no different from that of other processes using a physical absorption solvent.

TABLE 23
GAS SCRUBBING PROCESSES

PHYSICAL ABSORPTION

1. Selexol
2. Estasolvan
3. Fluor Solvent
4. Purisol (H_2S removal only)
5. Rectisol
6. Binax
7. Glycol Dehydration (H_2S removal only)
8. Chemsweet (H_2S removal only)

CHEMICAL ABSORPTION

1. MEA
2. DEA
3. Fluor Econamine (DGA)
4. Glycol Amine
5. Catacarb
6. Giammarco - Vetrocoke
7. Benfield

PHYSICAL/CHEMICAL ABSORPTION

1. Sulfinol
2. Alkazid

PHYSICAL ADSORPTION

1. Molecular Sieve
2. Reserve Synthetic Fuels (Getty)

DRY OXIDATION

1. Activated Carbon (H_2S removal only)
2. Claus (H_2S removal only)
3. Iron Oxide (H_2S removal only)

LIQUID OXIDATION

1. Ferrox
 2. Perox
 3. Stretford
 4. Takahox
 5. Townsend
 6. Ozone
-

The process was developed jointly by Institut Francais du Petrole of France, and Friedrich Uhde, G.m.b.H. of West Germany. As of 1974, the process was demonstrated in only two pilot plants and had not as yet been commercially applied.

FLUOR SOLVENT PROCESS. The absorbing agent in this process is an anhydrous organic compound, propylene carbonate. The proprietary process is licensed by Fluor Engineers and Constructors, Inc. of Los Angeles. Regeneration of the solvent can usually be accomplished by a let-down of pressure without the application of heat. The process is best suited for purifying gases with very high concentrations of carbon dioxide and hydrogen sulfide and is most cost-effective for gas streams greater than 20 million cubic feet per day. (Typical digester gas streams are less than this amount.)

PURISOL PROCESS. The absorbing agent in this process is N-methyl-pyrrolidone (NMP). The proprietary process is licensed by Lurgi Mineraloltechnik GmbH. The Purisol process is suitable for the selective removal of hydrogen sulfide in the presence of carbon dioxide since hydrogen sulfide has an exceptionally high solubility in NMP. Due to a rather complex flow scheme, however, the process is usually considered only for plants with high concentrations of contaminants in the gas stream.

RECTISOL PROCESS. The absorbing agent in this process is methanol. Due to the solubility of compounds such as hydrogen cyanide, benzene, and organic sulfur compounds in methanol, the Rectisol process is capable of removing most of the impurities in digester gas. Since the absorbed gases are readily released from the solvent when the

pressure is reduced, heat requirements for the solvent regeneration step is low.

Additionally, the cleaned gas has a very low water content.

The principal disadvantages of the Rectisol process are:

1. Complex flow scheme, usually making the process economical only for large scale application; and
2. Relatively high vaporization losses of the solvent, caused by the appreciable vapor pressure of methanol even at low temperatures.

WATER SCRUBBING PROCESS. Water scrubbing is the oldest process in existence. The absorbing agent in this process is water. Regeneration of the solution is achieved by a release in pressure. Some of the power released in the pressure-reduction of the liquid from the absorber and subsequent expansion of the absorbed gas can be reclaimed. Power-recovery is a turbine coupled to the motor of the water pump used in returning the regenerated water to the absorber column. The main advantage of using water as an absorbent as compared to the other chemical absorption processes are:

1. Simple plant design (no heat exchangers or reboilers);
2. No heat load;
3. Inexpensive solvent;
4. Solvent is reactive with CO_2 and H_2S at low pressure (15 psi);
5. No nitrogenous solvent vapors introduced into the gas stream; and
6. Pipeline quality gas can be achieved.

The principal disadvantages of the water scrubbing process are:

1. Substantial loss of valuable constituents of the gas stream. Components such as hydrogen and methane will partially dissolve in the water along with the CO_2 and it is usually not economically feasible to recover these gases;

2. Very high pumping load due to the large amount of water handled;
3. CO₂ removal requires additional pumping; and
4. It is difficult to recover CO₂.

The most common water scrubbing process is the Binax System.

BINAX SYSTEM PROCESS. The Binax System uses water under pressure to remove carbon dioxide and hydrogen sulfide. Water is compressed and injected into the top of a pressurized tower (Scrubber). Raw gas is injected at the base of the tower and flows upward, counter-current to the flow of the water. As the gas and water mix, the water absorbs the contaminants. The water from the bottom of the scrubber tower is depressurized and piped to the top of a second tower for regeneration. In the regenerator tower the carbon dioxide and hydrogen sulfide are flashed from the water and vented. The water at the base of the regeneration tower is pumped back to the top of the scrubber tower and recirculated. The Binax system is very compatible with the wastewater treatment scheme and can even use secondary effluent as the absorbing solvent.

GLYCOL DEHYDRATION PROCESS. This process is capable of removing high concentrations of hydrogen sulfide, carbon dioxide, and moisture. The two most commonly used glycols for gas dehydration are diethylene glycol (DEG) and triethylene glycol (TEC). The advantages of glycols include unusual hygroscopicity (ability to retain moisture), good stability against heat and chemical decomposition, low vapor pressures, and it is readily available at moderate cost. In a typical glycol dehydration unit, water vapor is continuously absorbed from the process gas stream by countercurrent contact with a highly concentrated

glycol stream (95 to over 99 percent) in a packed or bubbly tray column. Regeneration of the rich glycol solution is usually accomplished by inert gas stripping with the application of a high level of heat. When maximum dehydration is required in large installations, vacuum regeneration can also be used. Two processes with wide commercial application are the Drizo dehydration process licensed by the Dow Chemical Company and the high dew point depression dehydration process licensed by Black, Sivalls & Gryson, Inc.

CHEMSWEET SYSTEM PROCESS. The Chemsweet process developed by C-E Natco incorporates the use of inorganic zinc and water to form a slurry for the removal of hydrogen sulfide. A vessel is used to contain the slurry. Raw gas under pressure enters the bottom, passes through the slurry and exits at the top of the vessel. As the gas contacts the zinc oxide (ZnO), two reactions occur; formation of zinc sulfide (ZnS) and formation of zinc carbonate ($ZnCO_3$). Disposal of the spent slurry is by trucking to an approved landfill site.

The process is similar to the Binax systems except that the absorbing solvent is a water slurry which is actually easier to recharge than other solvents. This keeps the overall costs of the process down. The main disadvantage is that an additional process is needed for carbon dioxide removal.

CHEMICAL ABSORPTION PROCESSES

MEA SCRUBBING PROCESS. The MEA scrubbing process is one of the most widely used systems for the removal of carbon dioxide and hydrogen sulfide. The absorbing agent in this process is an aqueous solution of the chemical absorbent monoethanolamine (MEA). Advantages of the MEA process include:

1. High reactivity and solution capacity at moderate concentrations;
2. High alkalinity;
3. Low solvent cost;
4. Good stability; and
5. Ease with which the absorbent can be reclaimed from contaminated solutions.

Inherent disadvantages of the process are:

1. High vaporization losses; and
2. The MEA's formation of irreversible reaction products with impurities such as carbon-oxygen-sulfur (COS) and cesium (CS_2), resulting in excessive product losses if the gas contains significant amounts of these compounds.

Monoethanolamine solutions are highly corrosive at concentrations exceeding about 20 percent by weight. This could occur when heavily loaded with digester gas. Another drawback is the high heat requirement for regeneration of the solvent. Typical solution concentration used for the MEA process is about 20 to 25 percent by weight. Since it has been used almost exclusively for many years for the removal of hydrogen sulfide and carbon dioxide from natural gas, most other newer processes are typically compared with the MEA process.

DEA SCRUBBING PROCESS. The absorbing agent in this process is diethanolamine (DEA). It is the second most widely used gas treating process for hydrogen sulfide and carbon dioxide removal. Advantages of this process over MEA are:

1. Significantly lower vaporization losses; and
2. Low reactivity with COS and CS_2 .

Diethanolamine solutions are also generally less corrosive than monoethanolamine solutions. Disadvantages when compared to MEA include:

1. Lower reactivity; and
2. Higher solvent costs.

Additionally, reclaiming of contaminated solutions may require vacuum distillation. One version of the process, known as the S.N.P.A. DEA process, uses a relatively concentrated aqueous diethanolamine solution (20 to 30 percent by weight) and is capable of producing a purified gas satisfying pipeline specifications. Decomposition products which may be formed are removed quite easily by filtration through activated carbon. Sometimes, triethanolamine (TEA) is used, although it has been largely displaced because of:

1. Its low capacity (resulting from higher equivalent weight);
2. Its low reactivity; and
3. Its relatively poor stability.

FLUOR ECONAMINE PROCESS. The absorbing agent in this process is an aqueous solution of Diglycolamine (DGA). This compound couples the stability and reactivity of monoethanolamine with the low vapor pressure of diethylene glycol and can be used in more concentrated solutions (typically 40 to 70 percent by weight) than MEA; thereby, permitting reduced circulation rates and steam consumption. This results in substantial savings in both capital and operating costs. The process scheme is identical to a MEA system. Disadvantages of DGA are its relatively high maintenance cost due to its high corrosivity toward carbon steel.

GLYCOL-AMINE PROCESS. The absorbing agent in this process is a mixture of monoethanol-amine with di- or triethylene glycol and can be used for simultaneous digester gas scrubbing and dehydration. Major advantages of the process are the feature of simultaneous purification,

dehydration and somewhat lower steam consumption when compared to aqueous systems. Furthermore, the solvent can be stripped almost completely of hydrogen sulfide and carbon dioxide, resulting in the capability of producing extremely high purity treated gas. The principal drawbacks include the requirement that water content of the solution be kept at or below 5 percent in order to be effective as a dehydrating agent. This requires relatively high reboiler temperatures at which severe corrosion arises, making the utilization of expensive corrosion-resistant construction materials a necessity. The glycolamine process also has a high vaporization loss of the solvent.

CATACARB PROCESS. The Catacarb process, developed by Eickmeyer uses an absorbing agent containing a potassium salt and a nontoxic, stable, corrosion inhibiting additive. The process can remove hydrogen sulfide and carbon dioxide. The proprietary process has been installed at over eighty industrial plant locations and is licensed by Eickmeyer & Associates of Overland Park, Kansas. The principal advantages of the Catacarb process are high treated gas purity, lower energy consumption than required for hot potassium carbonate or monoethanolamine, and smaller equipment. The result is a savings in capital investment when compared to the conventional MEA process. The principle disadvantage is the difficulty in disposing of the potassium salt. A Class I dump site is required. Although several installations exist, little data has been available regarding this process.

GIAMMARCO-VETROCOKE PROCESS. The absorbing agent in this process for hydrogen sulfide and carbon dioxide removal involves the use of aqueous sodium or potassium carbonate solutions to which essentially stoichiometric proportions of arsenic trioxide are added. The process

is licensed by Giammarco-Vetrocoke of Marghera, Italy. The use of arsenites increases the absorption rate and capacity by two to three times those of other hot carbonate processes. The arsenite solution of the Giammarco-Vetrocoke process is regenerated by heat. The corrosion inhibiting nature of arsenic salts results in less stringent requirements for process unit construction materials. Economic comparisons of the Giammarco-Vetrocoke process with the hot potassium carbonate and ethanolamine process by Riesenfeld and Mallowney and Jenett show a definite economic advantage of the former process over the other two processes with respect to both capital investment and operating costs. However, Riesenfeld and Kohl have pointed out that those comparisons do not take into account the more recently developed activated potassium carbonate solutions and design improvements in the ethanolamine process. Although several hundred plants using this process are in operation world-wide, it has found little acceptance in the United States; probably because of the toxicity of the arsenite solution which requires special precautions in handling.

BENFIELD PROCESS. This process, originally developed by the U.S. Bureau of Mines, uses hot potassium carbonate solution as the absorbing agent. Subsequent improvement of the process by the use of activating additives in the solution was developed by Benson and Field of the Benfield Corporation. Carbon dioxide and hydrogen sulfide are removed in the absorber at elevated temperature (240 degrees F) and pressure (100 to 2,000 psig). Regeneration of the scrubbing solution takes place at about atmospheric pressure. Because of the high absorber temperature, the steam which other regenerative processes require to heat the solution to stripping temperature is not required. This

results in less regeneration heat requirement. The process can achieve high purity grade product gas and negligible methane losses. However, the process can be corrosive and bicarbonate crystal formation occurs if the solution is cooled at any point in the circuit. Foam inhibitors must also be added in some cases to avoid column flooding and dumping. An improved version of the Benfield process, known as the HiPure process, has been developed recently. The HiPure process uses two independent but compatible circulating solutions in series to achieve a high purity gas combined with high efficiency. Regeneration heat requirement is reduced by more than 20 percent. However, a 5 to 10 percent increase in capital investment cost is required. Thermal efficiency of the process is around 86 percent and operating cost is 75 percent of that required for a MEA process.

PHYSICAL/CHEMICAL ABSORPTION PROCESSES

SULFINOL PROCESS. This process is unique in that it employs a mixture of a chemical and physical solvent as the absorbing agent for removing hydrogen sulfide. The solvent consists of sulfolane (tetrahydrothiophane dioxide), diisopropanolamine (DIPA), and water. The proprietary process is licensed by Shell International Research Mij, N.V.; The Hague, The Netherlands; and Shell Development Co., Houston. The sulfolane acts as the physical solvent while DIPA acts as the chemical solvent. The ratio of each component is determined by the application specifications. The presence of the physical solvent enhances the solution capacity appreciably over that of a conventional amine process, especially at high operating pressures, resulting in lower circulation rates. The Sulfinol solvent is stable from degradation and is not very

corrosive. Regeneration of the solvent is accomplished by flashing followed by stripping at elevated temperatures. In addition to scrubbing digester gas to pipeline specifications, the Sulfinol process is capable of effectively removing carbonyl sulfide and mercaptans. Over 100 units of the process are in operation or under construction. The problem arises in trying to maintain the ratio of the physical/chemical absorbent. It has a very small range and, if it is exceeded, the efficiency of the process is significantly reduced.

ALKAZID PROCESS. The alkazid process, developed by I.G. Farben Industries in Germany, uses three different types of solutions as absorbing agents:

1. Alkazid "M" (an aqueous solution of potassium salt of methyl amino propionic acid) for absorbing either H_2S or CO_2 when present alone, or for absorbing both gases simultaneously;
2. Alkazid "DIK" (a potassium salt of diethyl or dimethyl amino acetic acid) for the selective removal of hydrogen sulfide; and
3. Alkazid "S" (a solution of sodium phenolate) for treatment of gases containing an appreciable amount of other impurities such as HCN, ammonia, carbon disulfide, mercaptans, dust, and tar. However, the process using this solution has not been commercialized.

The absorption solutions of the Alkazid process are regenerated by heat. The process is suitable for treatment of gases at atmospheric and elevated pressures for removing both hydrogen sulfide and carbon dioxide. Alkazid solutions are relatively noncorrosive and have low vapor pressure resulting in no solution losses. The solubility of hydrocarbons in alkazid solutions is low, resulting in minimum loss of hydrocarbons. Although this is an advantage over physical solvent processes, the ratio of the three absorbents is critical and small

variations can result in significant reductions in efficiency. Manufacturers of the process include Davy Powergas.

PHYSICAL ADSORPTION PROCESSES

MOLECULAR SIEVE PROCESS. Molecular sieves function by adsorbing small molecules while excluding large ones so that separations can be made based on molecular-size differences. Several types of molecular sieves capable of selective removal of hydrogen sulfide and carbon dioxide have been developed. Selection of the appropriate type depends on the nature and quantity of impurities to be removed. Sieve life ranges from 2 to 5 years for hydrogen sulfide and carbon dioxide removal, and 3 to 7 years for dehydration.

Regeneration is accomplished by purging under vacuum and flushing by the product gas. Molecular sieves have the advantage that they are conveniently skid-mounted and suitable for processing gas rates as low as a few thousand standard cubic feet per day. Typically, a minimum of three sieves are needed to complete the gas cleaning cycle. This makes the capital costs very high in comparison to the other alternatives.

There was some indication during our telephone survey that research is being done in Mountain View, California using molecular sieves for carbon dioxide removal of landfill gas. Costs for labor and management and a study of corrosion problems should be available soon.

RESERVE SYNTHETIC FUELS, INC. (GETTY SYNTHETIC FUELS) PROCESS.

This process was developed by Reserve Synthetic Fuels, Inc., which is the owner and operator of the first plant in the world to recover methane gas from sanitary landfills. The gas is first compressed and

passed through a small tower packed with activated carbon to remove hydrocarbons, hydrogen sulfide and water. The gas is then passed under higher pressure through a larger tower filled with a molecular sieve called zeolite for the removal of carbon dioxide. Regeneration is accomplished by heat in the first tower and by vacuum in the second. Regeneration of the activated carbon has a high heat requirement pushing the operating costs above other processes. Additionally, molecular sieves are used in conjunction with this process which further increases the costs.

A similar process is undergoing extensive testing at General Electric but the technology and economics are not completed. Their preliminary conclusions are that costs will be one-half to one-third that of similar systems.

DRY OXIDATION PROCESSES

INTRODUCTION. Most dry oxidation processes provide simple and effective removal of hydrogen sulfide only from gas streams. It is most commonly used where air quality control is strictly enforced. The non-regenerative nature of the supportive materials for the oxygen carriers necessitates regular recharging procedures which often require additional ground-space and labor. Since low purity of the recovered elemental sulfur can be a drawback of many dry oxidation processes, many regenerative processes have been developed which dissolve or suspend the oxygen carriers necessary for the oxidation of hydrogen sulfide in aqueous media. These processes usually allow continuous operation and are capable of recovering elemental sulfur of high purity.

ACTIVATED CARBON PROCESS. This process takes advantage of the catalytic action of activated carbon in the oxidation of hydrogen

sulfide to elemental sulfur at ordinary temperatures. The sulfur deposited on the activated carbon is recovered by extraction with ammonium sulfide. The main advantage of the activated-carbon process is that very pure sulfur is obtained in a relatively simple operation. The main drawback lies in the fact that the carbon can be deactivated rather rapidly by deposition of tar and polymeric materials on the particle surfaces, but this is unlikely to occur with digester gas. Costs for regenerating activated carbon are high, as indicated in the Reserve Synthetic Fuels Process.

CLAUS PROCESS. The Claus process should be considered more as a sulfur recovery process than as a gas purification process. The process recovers elemental sulfur from pure gaseous hydrogen sulfide or from streams containing hydrogen sulfide in high concentrations. Residual sulfur compounds are incinerated to sulfur dioxide. The Claus process yields sulfur of high purity and is widely used in the production of that element. Since the concentration of hydrogen sulfide has to be in sufficient concentration to sustain continuous combustion, the Claus process is usually used only when the gas stream has a fairly high hydrogen sulfide concentration. Carbon dioxide has to be removed by an additional process.

The addition of a dehydration unit for carbon dioxide and moisture removal results in a well cleaned digester gas. The costs may be competitive with other single processes which provide the same results.

IRON OXIDE PROCESS. The iron oxide process, also known as the iron-sponge process, is one of the oldest methods used in the removal of sulfur compounds from gas streams. It is also one of the simplest and

most commonly used processes in existing wastewater plants. In this process, hydrogen sulfide is removed completely by reaction with hydrated ferric oxide to form ferric sulfide. The ferric oxide is usually impregnated on wood shavings which serve as an inexpensive carrier. The sponge bed can be periodically or continuously regenerated by introduction of atmospheric oxygen which oxidizes ferric sulfide to elemental sulfur and ferric oxide. Accumulation of the sulfur can cover the surface of the oxide and cause loss of activity of the sponge and excessive pressure drop through the bed. Therefore, the container has to be recharged occasionally.

The iron oxide process does not remove nonoxidizable agents such as carbon dioxide. To accomplish this, a follow-up process, either molecular sieves or dehydration, must be provided. The cost for this combination can be very high.

Although many wastewater treatment plants have, in the past, selected the iron oxide process for hydrogen sulfide removal, few remain operating today. It appears that the biggest problem is regenerating the solvent.

In order to provide complete cost for gas cleaning with a dry oxidation process, a dehydration unit has to be included for the carbon dioxide removal. The Kemp dryer (Glen Burnie, Maryland) was selected along with a silica gel media produced by its subsidiary, Novatec. The solvent requires heat (500 degrees F) which can be provided by a heat exchanger with solvent replacement every 5 to 7 years. Carbon dioxide removal can be as high as 93 percent. High levels of moisture removal will also be provided.

LIQUID OXIDATION PROCESSES

INTRODUCTION. Liquid phase oxidation processes are primarily used for the treatment of gases containing relatively low concentrations of hydrogen sulfide in the presence of substantial concentrations of carbon dioxide. The basic chemical mechanisms in liquid oxidation processes are similar to those occurring in dry oxidation processes.

FERROX PROCESS. The Ferrox process, developed by the Koppers Company, uses a solution normally containing 3 percent sodium carbonate and 0.5 percent ferric hydroxide. The reaction of ferric hydroxide with hydrogen sulfide in the absorber is followed by the conversion to elemental sulfur by contact of the solution with air in the regenerator. The Ferrox process reportedly requires less ground area and lower labor and initial installation costs than conventional dry-box purification processes. However, complete removal of hydrogen sulfide often cannot be obtained as readily. Most Ferrox plants have recently been replaced by improved, more efficient processes.

PEROX PROCESS. This process uses an aqueous ammonia solution containing 0.3 grams per liter of an organic oxidation catalyst, usually hydroquinone, to absorb hydrogen sulfide. Recovery is accomplished by oxidation of ammonium hydrosulfide to elemental sulfur by contact with air. The main advantages of the process are nontoxicity of the solution used and slow formation of undesirable side-reaction products.

STRETFORD PROCESS. This process utilizes an aqueous solution containing sodium carbonate and bicarbonate, sodium vanadium, and sodium salts of anthraquinone disulfonic acid (ADA). Hydrogen sulfide is absorbed by the alkali solution and is removed through oxidation by vanadium into elemental sulfur. The proprietary Stretford process is

widely used in Europe and is licensed by British Gas Corporation - International Consultancy Service.

TAKAHAX PROCESS. The Takahax process utilizes naphthoquinone compounds dissolved in aqueous alkaline, usually sodium carbonate, solution as the oxygen carrier to oxidize hydrogen sulfide to sodium bisulfide and sodium bicarbonate. Oxidation of the bisulfide to elemental sulfur occurs in the absorber instantaneously. The proprietary process was developed and licensed by Tokyo Gas Company, Ltd.

TOWNSEND PROCESS. The Townsend process utilizes a concentrated stream of di- or triethylene glycol (typically 98 percent glycol and 2 percent water) containing dissolved sulfur dioxide. With water acting as the catalyst, hydrogen sulfide is reacted rapidly with sulfur dioxide to form elemental sulfur which can be recovered later by settling. The treated gas leaving the absorber is washed in a high-pressure column with concentrated glycol to remove sulfur dioxide carried by the gas out of the reactor. Like the above processes, the Townsend Process is primarily one of sulfur recovery. Furthermore, it is not as well-known as the Claus Process.

OZONE SCRUBBING SYSTEM. For many years, Ozone has been used in Europe for the destruction of hydrogen sulfide and carbon dioxide in controlling odors. Recently, advancements both in the application and production of ozone have greatly reduced the cost and increased the efficiency of the system.

The ozone scrubbing system employs the use of a mist scrubber with reaction chamber and a demister vain unit. Water is used as the removal medium and collection of the contaminants for removal from the gas stream. The water is adjusted to a pH of 13.5 or higher before it

is sprayed into the gas stream. The spray is of a very fine type or fog and once in the gas stream reacts very quickly with the contaminants to contain and remove them. The sprayed mist is then recaptured at the end of the reaction chamber and removed through the use of the demister vain unit. The water then travels by gravity to another tank where it is mixed with the ozone gas. This tank is known as a contact tank and is sized to provide a one minute detention time for mixing with the ozone gas. From the contact tank the water flows into a reaction tank where any residual ozone is allowed to react with any remaining contaminant and solids are settled out. From this point the water is pumped back to the spray scrubber and the pH is readjusted in route.

Although the ozone scrubbing process is used in Europe, there is little, if any, use or interest for its use, in the United States.

COMPARISON OF THE GAS SCRUBBING PROCESSES

Not all of the 28 previously described gas scrubbing processes are suitable to wastewater produced digester gas. A more detailed analysis will be made for those systems capable of removing hydrogen sulfide and carbon dioxide and where the process has a history of successful operation either in the wastewater industry or on gases similar in chemical quality to anaerobic digester gas.

Table 24 compares the 28 gas scrubbing processes. The first step in selecting a process was to insure that it's operability was good. That is, the process had to be an established and successful one for gas scrubbing. The following processes were considered to have good operability:

TABLE 24
COMPARISON OF GAS SCRUBBING PROCESSES

	PHYSICAL ABSORPTION PROCESSES								CHEMICAL ABSORPTION PROCESSES								PHYS/CHEM ABSORPTION	PHYSICAL ADSORPTION	DRY OXIDATION PROCESSES				LIQUID OXIDATION PROCESSES						
	SELEXOL	ESTASOLVAN	FLUOR SOLVENT	PURISOL	RECTISOL	BINAX	GLYCOL DEHYDRATION	CHEMSWEET	MEA	DEA	FLUOR ECONAMINE	GLYCOL AMINE	CATACARB	CIANTARGO- VETROCOKE	BENFIELD	SULFINOL	ALKAZID	MOLECULAR SIEVE	RESERVE SYNTHETIC	ACTIVATED CARBON	CLAUS	IRON OXIDE	FERROX	PEROX	STRETFORD	TAKAHAY	TOWNSEND	OZONE	
OPERABILITY	GOOD	POOR	FAIR	FAIR	GOOD	GOOD	FAIR	GOOD	GOOD	GOOD	GOOD	GOOD	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD	FAIR	FAIR	GOOD	GOOD	FAIR	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD
QUALITY OF H ₂ S REMOVAL	H	M	H	H	H	H	H	H	H	H	H	H	H	H	H	M	H	H	H	H	H	H	H	H	H	H	H	H	H
QUALITY OF CO ₂ REMOVAL	H	M	H	H	H	H	H	L	H	H	H	H	H	H	H	M	H	H	N.A.	N.A.	L	L	N.A.	N.A.	N.A.	N.A.	N.A.	N.A.	H
SULFUR RECOVERY NEEDED FOR COST EFFECTIVENESS	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	YES	YES	NO	YES	YES	YES	YES	YES	YES	NO
SUITED FOR WASTE- WATER TREATMENT	GOOD	N.A.	POOR	POOR	FAIR	GOOD	FAIR	NA	GOOD	FAIR	GOOD	GOOD	POOR	POOR	POOR	N.A.	N.A.	GOOD	GOOD	FAIR	FAIR	GOOD	FAIR	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD
REMOVES MOISTURE	NO	NO	NO	NO	YES	NO	YES	NO	NO	NO	NO	YES	NO	NO	NO	NO	NO	YES	YES	NO	NO	NO	NO	NO	NO	NO	NO	NO	YES
SOLVENT CORROSIVE	NO	N.A.	NO	N.A.	N.A.	NO	NO	NO	YES	YES	YES	NO	NO	YES	YES	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO	NO
RELIABILITY OF PROCESS	GOOD	N.A.	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD	GOOD	GOOD	GOOD	GOOD	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD	FAIR	POOR	GOOD	GOOD	FAIR	FAIR	FAIR	FAIR	FAIR	FAIR	GOOD
CAPITAL COST	H	N.A.	H	H	H	H	M	M	M+	M+	M	M	M	M	H	H	H	H	H	M	H	L	M	H	N.A.	N.A.	N.A.	M	
OPERATING COST	H	N.A.	M	H	H	M	H	LOW	H	H	M	H	M	M	M	H	H	H	H	M	M	M+	H	H	N.A.	N.A.	N.A.	H	
MAINTENANCE COST	M	N.A.	M	H	H	M	M	L	H	H	H	H	H	H	H	H	H	H	H	H	M	M+	M	H	N.A.	N.A.	N.A.	H	
REGENERATION COST	H	N.A.	M	N.A.	M	M	H	M	M	N.A.	M	H	H	H	M	H	H	M	H	H	H	H	H	M	N.A.	N.A.	N.A.	M	

H - High
M - Moderate
L - Low
N.A. - Information Not Available

I. Physical Absorption

1. Selexol
2. Rectisol
3. Binax
4. Chemsweet

II. Chemical Absorption:

1. MEA
2. DEA
3. DGA (Fluor Econamine)
4. Glycol Amine

III. Physical Adsorption

1. Molecular Sieve

IV. Dry Oxidation

1. Claus
2. Iron Oxide (plus dehydration)

V. Liquid Oxidation

1. Ozonation

Of the physical absorption processes, only the Binax system was considered for further analysis. The other processes had much higher total costs.

Of the chemical absorption processes, the Glycol Amine process was not considered because its costs, particularly in solvent regeneration, were higher than any other chemical absorption process. The DGA process was also eliminated because of its corrosiveness and high costs. Otherwise the process is similar to both MEA and DEA, which will be further analyzed.

Molecular sieves were not considered further because of their very high capital costs. Systems utilizing molecular sieves as an add-on process for carbon dioxide were eliminated because this caused a very large increase in capital costs.

Of the dry oxidation processes the Iron Oxide process was considered with the addition of a dehydration unit.

Ozonation, the liquid oxidation process, was not included because of its lack of experience in the United States market.

In summary, then, the systems that will be further analyzed are:

1. Binax,
2. MEA,
3. DEA, and
4. Iron Oxide with dehydration.

ANALYSIS OF THE MORE VIABLE GAS SCRUBBING PROCESSES

Cleaning or scrubbing the gas can improve its utilization with two major benefits. First, equipment that will use the gas can be of standard materials and manufacture, and second the Btu value of the gas is increased to 1,000 Btu per cubic foot or more, so less gas needs to be handled for the same amount of energy.

Of the processes available for cleaning digester gas, only four were selected for further evaluation. These include: The Iron Sponge Process, the Amine Processes, (DEA and MEA), and the Binax process. Each of the above named processes offers a distinct advantage and varies in both capital cost and operation and maintenance costs. These advantages, with a process description and schematic diagram, will be examined and compared in the following sections.

PROCESS DESCRIPTION

INTRODUCTION. The following process descriptions expand upon the previous section for those more viable processes. Although the

information is general in nature, a process schematic is provided for comparison purposes.

BINAX SYSTEM. The Binax System was developed in the mid 1970's by a group of engineers in Southern California for the explicit use by wastewater treatment facilities for the removal of hydrogen sulfide and carbon dioxide from digester gas.

The system incorporates the use of water as the scrubbing medium. The process water can be taken from two different sources which, in effect, defines the type of system being used.

The most publicized wastewater treatment plant using a Binax System for digester gas scrubbing is the pilot plant in Modesto, California, (209) 577-5300.

A similar unit has been proposed for Livermore, California, and should be in operation in 1983.

Binax System I. The Binax System I is a liquid absorption method which incorporates the use of a packed tower. Absorption of the carbon dioxide and hydrogen sulfide is accomplished under pressure with the digester gas being compressed and injected into the base of the tower. As the gas flows up the tower, a water solution flows downward or countercurrent through the trays or packing and absorbs the contaminants. This results in purifying the gas which exits the top of the tower. The contaminated solution exits through the base. (Figure 45).

The solution containing the contaminants is then sent to a second tower where the hydrogen sulfide and carbon dioxide are flushed out and removed, with the carbon dioxide being released to atmosphere and the hydrogen sulfide being sent to an iron sponge unit.

Sewage Gas Vehicle Fueling System

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

GAS SCRUBBING
BINAX SYSTEM I

System I incorporates a closed loop water system. Water continuously circulates and makeup water is added as evaporation occurs.

Binax System II. Binax System II uses the same basic equipment as in System I with the exception that the water scrubbing solution is a once-through type system and employs the use of secondary effluent from the wastewater treatment facility (Figure 46).

After separating the hydrogen sulfide and carbon dioxide from the digester gas, the solution containing the contaminants is pumped directly back into the aeration basin where the contaminants are treated again.

Both systems offer a new and viable treatment process for removing hydrogen sulfide and carbon dioxide. Nonetheless, only two installations have been completed and in operation to date. Results from these two facilities have indicated some mechanical problems but overall the process has performed as designed and the contaminant removals have been better than expected.

Operation and maintenance procedures appear well suited to those found in a wastewater treatment plant and there is less reluctance to adopt this process to wastewater treatment.

AMINE PROCESS (MEA, DEA). The Amine Gas Sweetening Process is a widely used and accepted process in the natural gas industry. Monoethanol Amine (MEA) has been used exclusively for the removal of both hydrogen sulfide and carbon dioxide until recently when Diethanol Amine (DEA) was developed. Now, both processes are equally popular. The selection of the Amine Process to be used is based on the amount of contaminants in the gas. Both processes basically use the same equipment.

Sewage Gas Vehicle Fueling System

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

GAS SCRUBBING BINAX SYSTEM II

FIGURE 46

Following is a description of the Amine Process in general. The differences in design calculations between MEA and DEA are identified.

The digester gas enters the system and is contacted in a counter-current gas-liquid absorption process in a trayed or packed column or tower. This provides the gas with intimate contact for the chemical reaction between the hydrogen sulfide, carbon dioxide and the Amine (Figure 47). Past design practice dictates a scrubber on the inlet gas to remove any entrained liquids including water and distillate from the gas before it enters the column or tower. Additionally, a scrubber is required on the outlet gas to scour any Amine solution carried over from the contactor.

Rich Amine, from the bottom of the contactor, is flushed at reduced pressure to remove any entrained methane gases, and is heated in a rich-lean solution Amine exchanger. The solution from the exchanger is then fed to a stripping tower where the solution is regenerated and any residual methane gas is stripped by steam.

The gases that are stripped are concentrated in an overhead accumulator and disposed of by burning, either in a flare or a reboiler, or other type of incineration device. In cases where the sulfur content exceeds 3 to 5 tons per day, the acid gases may be processed for sulfur recovery.

Lean Amine solution from the bottom of the reboiler is exchanged with the rich Amine in the solution exchanger, pumped back to the contactor to complete and restart the process loop.

Since a clean Amine solution is the basic key to the success of an Amine treatment system, good filtration should be provided.

AMINE SCRUBBING SYSTEM

FIGURE 47

Design recommends granular activated carbon filtration, though these types of filters can be costly and a high maintenance item.

Monoethanolamine (MEA) is a primary amine normally used in a 15 to 20 percent solution in water. Generally, MEA is best used in systems where gas pressure is low and MEA solutions can be reclaimed in a side stream reclaimer.

Diethanolamine (DEA) is a secondary amine normally used in a 30 to 35 percent solution in water. It is slightly selective toward hydrogen sulfide when the carbon dioxide/hydrogen sulfide ratio in the gas stream is high. Removal efficiency of the contaminants will increase as the digester gas partial pressure in the stream increases. But, DEA cannot be reclaimed due to the high boiling point of the solution.

The Amine process is one of the most popular processes used in the oil and gas industries. Facilities utilizing both processes are found at the Hunt Oil Refinery in Longview, Texas and Bird and Sons Refinery in Martinez, California. Its use, however, in the wastewater treatment field may be limited due to both capital and operation and maintenance costs. Operating labor for the facility has to be dedicated to the process and will not have the availability of multi-purpose operations throughout the whole treatment facility.

IRON OXIDE (SPONGE) PROCESS. The Iron Oxide (Sponge) Process is a treatment unit containing an open vessel sized according to the volume of gas to be processed. (Figure 48) A screen or coarse packing is placed in the bottom of the vessel to support the sponge media. The remainder of the vessel is filled with the media leaving a small void space at the top. The sponge media consists of oak wood shavings

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**IRON SPONGE
SCRUBBING SYSTEM**

FIGURE 48

impregnated with an hydrated iron oxide. Digester gas enters the top connection and passes downward through the bed and exits out the bottom. Hydrogen sulfide reacts with iron oxide and forms iron sulfide.

The Iron Sponge Process is basically a batch process and requires a multiple tower system of several units for continuous operation. The outlet gas then enters a dehydration unit for absorption of the carbon dioxide and moisture.

Several wastewater plants across the country have installed the Iron Oxide process but few remain in operation. Various operating and maintenance problems have resulted in most of these installations to be abandoned. For additional information contact:

Fort Worth, Texas - (017) 277-7591

South San Francisco, California - (415) 877-8555

Regeneration of the sponge can be accomplished by the use of an air blower and bleeder valve. The air blower continuously circulates the gases contained within the tower while the bleeder valve allows 8 percent oxygen to be introduced into the system over a 24-hour period. Once regeneration is complete the bed-life of the media will be approximately 60 percent of the original. The system may be regenerated only twice before new media is required. Recharging of the media is both time consuming and labor intensive. The media has to be removed by hand and replaced the same way.

While the system may be used on low pressure gases, the efficiency is greater at higher pressures due to the reduced possibilities of short-circuiting by side wall channeling.

ECONOMIC ANALYSIS

As the cost of fuels continues to rise, digester gas utilization throughout the treatment plant and around a municipality becomes more attractive and cost-effective.

Previously, the establishment of the BTU value of digester gas had been described as; unscrubbed digester gas, 590 to 650 BTUs per cubic foot and scrubbed digester gas, 980 to 1050 BTUs per cubic foot. This indicates a differential of 370 to 400 BTUs per cubic foot or 400,000 BTUs per 1,000 cubic feet produced. Consequently, 4 therms of additional energy is available per 1,000 cubic feet of digester gas. (A "therm" equals 100,000 BTUs of energy.) Additionally, when the digester gas is in the higher BTU valve, it may be used in a variety of places or more readily sold to a public utility. Nonetheless, the cost associated with its conversion into a usable energy source can be substantial.

Internal Combustion engines using digester gas as a fuel source, have been in operation for years. These engines, in order to compensate for the impurities contained in raw gases, are derated or oversized and are not operated at the maximum capacity. In other uses where the raw gas is burned as a fuel for boilers and heating, dual fuel sources must be made available.

In choosing a system for gas scrubbing, a number of variables must first be examined.

- 1) Quantity of gas to be cleaned?
- 2) How much Carbon Dioxide (CO_2) and Hydrogen Sulfide (H_2S) is contained in the gas?
- 3) What is the source of end use of the product?
- 4) Is the payback less than half of the remaining life expectancy of the treatment facility?

These four questions require a clearly defined answer before any type of system selection process can be started. For example:

Question one is important because the cost of the gas scrubbing system is dependent on the volume of the gas to be cleaned.

Question two is especially important in large systems to determine whether or not auxiliary sulfur recovery equipment is required.

In developing the answer to question three, the analysis for question four is started. In this analysis it is necessary to remember that the pay back in capital and operating cost should be less than one half the life expectancy of the plant because it has been established that at least one major overhaul to the equipment will take place in the first 5 to 8 years after being placed into operation.

Table 25 repeats those processes from Table 24 that were selected as the more viable alternatives for digester gas scrubbing.

The operability of each process is rated as good, which means that each is a tried and true method for gas cleaning. Each has a high level of hydrogen sulfide removal with the Binax process having a slight edge. The Iron Oxide process requires an additional process for carbon dioxide removal.

The costs are updated to an ENR index of 4500 as explained in the Economic Analysis in Chapter III. The capital costs and operation and maintenance costs are discussed for each process.

In determining these costs, the level of treatment for each process considered is "pipeline quality," except for moisture removal. That is, 10 ppm of hydrogen sulfide and 2 percent (volume) of carbon dioxide in the final product.

TABLE 25

COMPARISON OF THE MORE VIABLE
GAS SCRUBBING PROCESSES

	PHYSICAL ABSORPTION	CHEMICAL ABSORPTION		DRY OXIDATION
	BINAX I & II	MEA	DEA	IRON OXIDE
OPERABILITY	GOOD	GOOD	GOOD	GOOD
QUALITY OF H ₂ S REMOVAL	HIGH	HIGH	HIGH	HIGH
QUALITY OF CO ₂ REMOVAL	HIGH	HIGH	HIGH	REQUIRES DEHYDRA- TION
SULFUR RECOVERY NEEDED FOR COST- EFFECTIVENESS	NO	NO	NO	NO
SUITED FOR WASTE- WATER TREATMENT	GOOD	GOOD	FAIR	GOOD
REMOVES MOISTURE	NO	NO	NO	YES
SOLVENT CORROSION	NO	YES	SOME	NO
RELIABILITY OF PROCESS	FAIR	GOOD	GOOD	GOOD
CAPITAL COST	HIGH	MOD+	MOD+	LOW
OPERATING COST	MOD	HIGH	HIGH	MOD+
MAINTENANCE COST	MOD	HIGH	HIGH	MOD+
REGENERATION COST	MOD	MOD	MOD	HIGH

Binax Systems. The Binax Systems offer a new concept in gas cleaning. Though capital costs are quite high (Figure 49), the operating and maintenance costs are low (Figure 50) and the systems seem to be readily adaptable to waste treatment facilities. Reviewing the two systems offered, Binax II is preferred because of the use of existing nonpotable water in the cleaning of the gas and returning the water to aeration.

Amine Systems. The Amine Systems (MEA, DEA) are the most costly processes (Figure 51). These systems remove both hydrogen sulfide and carbon dioxide. Operating costs include such items as chemical replacement, electrical and labor. Manpower operating the facility must be dedicated to the operation and constant attention is required (Figure 52). Filtration of the amine solution is required for best operation and carbon filters are suggested as the most economical. Use of these systems is suggested only when extremely high volumes of gas are to be treated. Processes using the Amine System and contemplating adding the Claus System for sulfur recovery should be examined only when the gas production exceeds 5 million cubic feet per day or when the sulfur content of the gas is 3 to 5 tons of recoverable sulfur.

Iron Sponge Systems. The capital cost associated with this system is low in comparison to other systems; even with the addition of a dryer for carbon dioxide removal (Figure 53). Operating cost and the associated maintenance are higher than the other systems due to the necessity to replace the media by hand (Figure 54). These costs do not include the silica gel needed for carbon dioxide. For example, the Novatec System, a subsidiary of Kemp, estimated the capital cost at about \$500,000 for treating 1.44 million cubic feet of gas per day.

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR GAS SCRUBBING
WITH THE BINAX SYSTEMS**

FIGURE 49

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**OPERATION AND MAINTENANCE COSTS
FOR BINAX SCRUBBING SYSTEMS**

FIGURE 50

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**COSTS FOR GAS SCRUBBING
WITH THE AMINE PROCESSES**

FIGURE 51

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**OPERATION AND MAINTENANCE COSTS
FOR THE AMINE SCRUBBING SYSTEMS**

FIGURE 52

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION

**COSTS FOR GAS SCRUBBING
WITH IRON SPONGE**

FIGURE 53

FEASIBILITY STUDY
FOR ALTERNATIVE FUELS PRODUCTION
**OPERATION AND MAINTENANCE COSTS
FOR IRON SPONGE SCRUBBING**

FIGURE 54

Problems associated with the system are the possibility of a spontaneous fire during the cleaning out of the old media and gas channeling within the vessel due to low operating pressures. Advantages of the system are that it is compatible to clean small streams as well as large.

PROCESS SELECTION

Finally, Table 26 is provided to show the effectiveness of each process at various feed gas volumes. The "Y" represents yes the process is economical to use at the given flow rate. The "N" represents the process is not effective and should not be considered at that volume.

For example, if a volume of 55,000 cubic feet of digester gas is to be cleaned, only the iron sponge and MEA processes should be considered.

APPLICATION TO EAST BAY MUNICIPAL UTILITIES DISTRICT

As presented in Chapter IV, 1.29 million cubic feet of digester gas was estimated as the average daily gas production at the EBMUD plant. Referring to Table 26, the processes that can effectively clean such a large volume of gas include; iron oxide, DEA, and Binax. The costs for each are presented below:

<u>Process</u>	<u>Capital Cost</u>	<u>Operation and Maintenance</u>
Iron Oxide	\$321,000	\$47,000/yr
DEA	\$725,000	\$99,000/yr
Binax II	\$750,000	\$ 1,000/yr

TABLE 26

SELECTION OF GAS SCRUBBING
SYSTEM BASED ON VOLUME TO BE TREATED

TREATMENT	VOLUME (1000 cu. ft./day)								
	0-20	21-50	51-75	76-100	100- 250	251- 500	501- 750	751- 1,000	1,000+
IRON OXIDE (with dehy- dration)	Y	Y	Y	Y	Y	Y	Y	Y	Y
MEA	N	N	Y	Y	Y	Y	N	N	N
DEA	N	N	N	Y	Y	Y	Y	Y	Y
BINAX #1	N	N	N	Y	Y	Y	Y	Y	Y
BINAX #2	N	N	N	N	Y	Y	Y	Y	Y

Determining the present worth of the Iron Oxide and Binax systems over 20 years at 7-3/8 percent interest, which includes replacement of the iron sponge media every 5 years, results in the following:

IRON OXIDE

Capital Cost	\$ 321,000
PW O&M	\$ 484,000
Total Present Worth	\$ 805,000
Average Annual Cost	\$ 78,000

BINAX II

Capital Cost	\$ 750,000
PW O&M	\$ 10,000
Total Present Worth	\$ 760,000
Average Annual Cost	\$ 74,000

Consequently, the Binax II System would be recommended. This, of course, is based on continuous treatment of all the 1.29 million cubic feet of digester gas. The design would essentially duplicate the schematic on Figure 46 and be fitted into the plant with the necessary inlet/outlet piping. The scrubbing water would be taken from the secondary clarifiers and the waste process water returned to the oxygenation tanks. The impact of this return process water was considered negligible as far as an added treatment cost to the plant.

LITERATURE REVIEW

LITERATURE REVIEW

A. SOLIDS MANAGEMENT

1. Culp, Wesner, and Culp. Cost and Performance Handbook, Sludge Handling Processes, Wastewater Treatment and Reuse Seminar. South Lake Tahoe. October, 1971.
2. Greifer, Lois. "Reuse, Energy, Sludge: In What Direction Are We Going?" Water and Wastes Engineering, Vol. 15, No. 1, 1978.
3. Jacobs, A.; Brailey, D.; and Pickart, B. "Municipal Sludge Management for Recovering Energy. Part 2" Water and Sewage Works, Vol. 126, No. 10, 1979.
4. Jacobs, A.; Brailey, D.; and Pickart, B. "State of the Art of Municipal Sludge Management. Part 1" Water and Sewage Works, Vol. 126, No. 8, 1979.
5. Matei, Peter F. "Sludge Management System for the Metropolitan St. Louis Sewer District" Proceeding of the National Conference on Municipal Sludge Management, June, 1974.
6. Riehl, M. L.; Weiser, H. H.; and Rheins, B. T. "Effects of Lime Treated Water on Survival of Bacteria" Journal of the American Water Works Association, Vol. 44, No. 5, 1952.
7. Sontheimer, H. "Effects of Sludge Conditioning With Lime on Dewatering" Proceedings of the Third International Conference (Munich), Advances in Water Pollution Research, Vol. 2, 1966.
8. Torpey, Wilbur. A Method of Reducing Sludge Produced From An Activated Sludge Plant. (Unpublished) Environmental Technology Consultants, Inc., Springfield, VA. January, 1981.
9. Tortovici, L. and Stahl, J. F. "Waste Activated Sludge Research" Proceedings of Sludge Management, Disposal, and Utilization Conference. Information Transfer Inc., Rockville, MD 20852. 1977.
10. USEPA. Process Design Manual for Sludge Treatment and Disposal. Technology Transfer. EPA 625/1-74-006. October, 1974.
11. USEPA. Process Design Manual for Sludge Treatment and Disposal. Technology Transfer. EPA 625/1-79-011. September, 1979.
12. USEPA. Sludge Handling and Conditioning. Office of Water Program Operations. Washington D.C. 20460. EPA 430/9-78-022. February, 1978.
13. Vesilind, P. Aarne. Treatment and Disposal of Wastewater Sludges. Ann Arbor Science Publishers, Inc., 1974.

B. ANAEROBIC DIGESTION: DESIGN AND OPERATION

1. Alberston, O. E. "Ammonia Nitrogen and the Anaerobic Environment" *Journal of the Water Pollution Control Federation*, Vol. 33, p. 978, 1961.
2. Barber, N. R. "Keep Your Digester in Good Shape" *Water and Wastes Engineering*, Vol. 14, No. 9, 1977.
3. Boarer, J. "Unusual Waffle Bottom Digester Design" Presented at the California Water Pollution Control Association, Northern Regional Conference. Redding, CA. October 20, 1978.
4. Buzzell, J. C. and Sawyer, C. M. "Biochemical vs. Physical Factors in Digester Failure" *Journal Water Pollution Control Federation*, Vol. 35, p. 205, 1963.
5. Collins, A. S. and Gilliland, B. E. "Control of Anaerobic Digestion Process" *Journal Environmental Engineering Division - ASCE*, Vol. 100, EE2, p. 487, 1974.
6. Dague, R. R. "Application of Digestion Theory to Digester Control" *Journal of the Water Pollution Control Federation*, Vol. 40, No. 12, 1968.
7. Davis, G. H. "New Twist in Digester Design" *American City and County*, p. 68, May, 1976.
8. Dietz, J. C.; Clinebell, P. W.; and Strub, A. C. "Design Considerations for Anaerobic Contact Systems" *Journal Water Pollution Control Federation*, Vol. 38, p. 517, 1966.
9. Drnevich, R. F. and Matsch, L. C. "A New Sludge Digestion Process" Fifth National Conference on Acceptable Sludge Disposal Techniques. January, 1978.
10. Envirotech Process Equipment. Digestion Equipment (Digester Covers, Digester Mixers) Consulting Engineering Design Manual.
11. Fuhrman, R. E. "Scum Control in Sludge Digestion" *Sewage and Industrial Wastes*, Vol. 26, No. 4, 1954.
12. Johnson, W. H. and Coombs, J. M. "Anaerobic Digestion Design Advances Increase Energy Production" (Unpublished).
13. Kotze, J. P. "Anaerobic Digestion - Characteristics and Control of Anaerobic Digestion" *Water Research*, Vol. 3, p. 545, 1969.
14. Lawrence, A. W. "Application of Process Kinetics to Design of Anaerobic Processes" Anaerobic Biological Treatment Process Advances in Chemistry. Series 105 American Chemical Society, Washington, DC 20036. 1971.

15. Lynam, B.; McDonnell, G.; and Krup, M. "Start-Up and Operation of Two New High-Rate Digestion Systems" *Journal of the Water Pollution Control Federation*, Vol. 39, No. 4, 1967.
16. Malina, J. "Variables Affecting Anaerobic Digestion" *Public Works*. September, 1962.
17. McCarty, P. L. and McKinney, R. E. "Salt Toxicity in Anaerobic Treatment" *Journal of the Water Pollution Control Federation*, Vol. 33, p. 399, 1961.
18. Mignone, Nicholas A. "Anaerobic Digestion Design for Energy Generation" *Public Works*, No. 10. 1974.
19. Mueller, L. E.; Hindin, E.; Lunsford, J. V.; and Dunstan, G. H. "Some Characteristics of Anaerobic Sludge Digestion, I. Effect of Loading" *Sewage and Industrial Wastes*, Vol. 31, p. 669, 1959.
20. McCarty, P. L. "Anaerobic Processes" Presented at the Birmingham Short Course on Design Aspects of Biological Treatment, International Association of Water Pollution Research, Birmingham, England, 1974.
21. McCarty, P. L. "Anaerobic Waste Treatment Fundamentals - Part 2 - Environmental Requirements and Control" *Public Works*, p 123. October, 1964.
22. McCarty, P. L. "Anaerobic Waste Treatment Fundamentals. Part 4 - Process Design" *Public Works*. December, 1964.
23. McCarty, P. L. "Kinetics of Waste Assimilation in Anaerobic Treatment" Development in Industrial Microbiology American Institute of Biological Sciences, No. 7, 1966.
24. Schroeffer, G. J. and Ziemke, N. R. "Development of the Anaerobic Contact Process" *Sewage and Industrial Wastes*, Vol. 31, No. 2, 1959.
25. Spencer, R. R.; Wong, A. L.; Coates, J. A.; and Ahlstrom, S. B. Operational Characteristics of Anaerobic Digesters at Selected Municipal Wastewater Treatment Facilities in the United States U.S. Department of Energy Contract EY-76-C-06-1830. Prepared by Pacific Northwest Laboratory, Batelle Memorial Institute, Richland, WA. December, 1978.
26. Tavery, M. A. and Nelson, J. "New Problems Can Occur With Anaerobic Digestion" *Water and Waste Engineering*. December, 1979.
27. Walker, J. "Successful Anaerobic Digestion - A Review" (Unpublished) Peabody-Wells Company. 1978.
28. Water Pollution Control Federation. Manual of Practice No. 16. Anaerobic Sludge Digestion Washington, DC 20016. 1968.

29. Woods, C. E. and Malina, V. F. "State Digestion of Wastewater Sludge"
Journal of the Water Pollution Control Federation, Vol. 37, p 1495,
1965.

C. SLUDGE THICKENING

1. Bird Machine Company Product Manual South Walpole, MA. February, 1971.
2. Bradney, L. and Bragstad, R. E. "Concentration of Activated Sludge by Centrifuge." Sewage and Industrial Waste, Vol. 27, p. 404, 1955.
3. Dick, R.I. "Thickening." Water Quality Engineering: New Concepts and Developments, E.L. Thackston and W.W. Eckenfelder, Jr. Editors, Jenkins Publishing Co., New York, 1972.
4. Dick, R.I. and Ewing, B.B. "Evaluation of Activated Sludge Thickening Theories." Journal of the Environmental Engineering Division. ASCE. Vol. 93, EE4, p. 9, 1967.
5. Gulas, V., Benefield, L. and Randall, C. "Factors Affecting The Design of Dissolved Air Flotation Systems." Journal of the Water Pollution Control Federation. Vol. 50, p. 1835. 1978.
6. Keninath, T.M., Ryckmon, M.D., Dana Jr. C.H., and Hofer, D.A. "Design and Operational Criteria for Thickening Biological Sludges Parts I, II, III, and IV." Water Resources Research Institute, Clemson University. September 1976.
7. Keith, F.W. and Little, T.H. "Centrifuges in Water and Waste Treatment." Chemical Engineering Progress. Vol. 65, November, 1969.
8. Komline, T.R. "Dissolved Air Flotation Tackles Sludge Thickening." Water & Wastes Engineer. p. 63, February, 1978.
9. Komline-Sanderson Engineering Corporation. Flotation Thickening Practice. Bulletin No. KSB-124-7403. Peapack, New Jersey.
10. Noland, R.F. and Dickerson, R.B. "Thickening of Sludge." USEPA Technology Transfer Seminar on Sludge Treatment and Disposal. Vol. 1 USEPA-MERL. Technology Transfer. Cincinnati, Ohio 45268. October 1978.
11. Torpey, W.N. "Concentration of Combined Primary and Activated Sludges in Separate Thickening Tanks." American Society of Civil Engineers. Volume 80, No. 443, 1954.
12. Walker Process Corporation. Thickeners For Liquid - Solids Separation. Bulletin No. 13-I-80. Copyright 1980. Walker Process Corporation.

D. ENERGY CONSERVATION

1. Budde, P.E., Doran, M.D. and Ratai, W.R. "Energy Recovery from Wastewater Treatment Plant Effluent." Journal of the Water Pollution Control Federation. Vol. 51, No. 8, 1979.
2. Chiu, Yishu and Guey-Lee, William. "Converting Wastes to Energy -A Conceptual Approach." National Science Foundation. February, 1978.
3. Foster, Walter E. "Conserve Energy in Wastewater Systems." Water and Wastes Engineering. Volume 15, No. 4, 1978.
4. Frostell, Bjorn. "Wastewater: Energy for the Future?" Water and Wastes Engineering. Volume 126, No. 6, 1979.
5. Jacobs, A. "Managing Energy at Water Pollution Control Facilities." Water and Sewage Works. Volume 127, No. 8, 1980.
6. Jones, J.L. and Bomberger, Jr. D.C. "Energy Usage and Recovery in Sludge Disposal." Part 1. Volume 124, No. 7 and No. 8, 1977.
7. Reid, Crowther and Partners, Ltd. "Designing an Energy Efficient Wastewater Treatment Plant." Water and Sewage Works. Part 1, Volume 126, No. 11. Part 2, Volume 126, No. 12. (1979) Part 3, Volume 127, No. 2. (1980).
8. Roberts, E.B. and Hagan, R.M. "Explore the Energy Economics of Wastewater Treatment and Reuse." The Bulletin. Volume 13, No. 1, 1976.
9. Rushbrook, Jr. E.L. and Wilke, D.A. "Energy Conservation and Alternative Energy Sources in Wastewater Treatment." Journal of the Water Pollution Control Federation. Volume 52, 1980.
10. Smith, James E. "Inventory of Energy Use in Wastewater Sludge Treatment and Disposal." Proceedings of the First Mid-American Conference on Environmental Design. Kansas City. May, 1976.
11. Smith, R. "Electrical Power Consumption for Municipal Wastewater Treatment." Environmental Protection Technology Service. EPA-42-73-281. July, 1973.
12. USEPA. "Electrical Power Consumption for Municipal Wastewater Treatment." Environmental Protection Technology Series. EPA-R2-73-281. July, 1973.
13. USEPA. "Energy Requirements for Municipal Pollution Control Facilities." EPA Technology Series. Cincinnati, Ohio 45268. EPA-600/2-77-214. November, 1977.
14. Wesner, G.M., Culp, G.L. Lineck, T.S. and Hinricks, D.J. "Retrofitting for Energy Savings." Water and Sewage Works. Volume 127, No. 6, 1980.

D. ENERGY CONSERVATION (Continued)

15. Winkler, W.W. and Welch, F.M. "Energy Conservation Dictates Innovative Treatment Plant Design." Public Works. March, 1974.
16. Young, R.A. "Energy Conservation: A By-Product of Efficient Sewage Treatment." Pollution Engineering. Volume 11, No. 5, 1979.

E. HIGH-RATE DIGESTER LOADING

1. Estrada, A.A. "Design and Costs Considerations in High Rate Sludge Digestion." Journal-Sanitary Engineering Division-ASCE. Volume 86, SA3, p. 111, 1960.
2. Sawyer, C.N. and Grumbling, J.S. "Fundamental Considerations in High Rate Digestion." Journal of ASCE. Volume 86, SA2, p. 49, 1960.
3. Zablatzky, A.R. and Baer, G.T. "High-Rate Digester Loadings." Journal of the Water Pollution Control Federation. Volume 43, P. 268, 1971.
4. Zoltak, J. and Gram, A.L. "High-Rate Digester Mixing Study Using Radioisotope Tracer." Journal of the Water Pollution Control Federation. Volume 47, p. 79, 1975.

F. DETENTION TIME IN DIGESTERS

1. Clark, R.H. and Orr, V.D. "Digestion: Concentration -Loading - Time Limits." Journal: Sanitary Engineering Division - ASCE. Volume 98, SA5, p. 809, 1972.
2. HinDrin, E. and Dunstan, G.H. "Effects of Detention Time on Anaerobic Digestion." Journal of the Water Pollution Control Federation. Volume 32, No. 9, 1960.

G. DIGESTER pH

1. Clark, R.H. and Speece, R.F. "The pH Tolerance of Anaerobic Digestion." Advances in Water Pollution Research. Volume 1. S.H. Jenkins, ed. Pergamon Press, Oxford, England. 1970.

H. DIGESTER TEMPERATURE

1. Cleary, E.J. "High and Low Temperature Digestion Experiments - Consecutive Digestion." Sewage Works Journal. Volume 7, p. 781, 1935.
2. Goleuke, C.G. "Temperature Effects on Anaerobic Digestion of Raw Sewage Sludge." Sewage and Industrial Wastes. Volume 0, p. 1225, 1958.
3. O'Rourke, J.T. Kinetics of Anaerobic Treatment at Reduced Temperatures. Unpublished doctoral dissertation. Stanford University. Palo Alto, California 94305. 1968.
4. Schwerin, D.J. "The Effect of Temperature on Anaerobic Digestion." unpublished master's thesis. Marquette University, Milwaukee, Wisconsin 53233. June, 1976.
5. USEPA. Use of Solar Energy to Heat Anaerobic Digesters. Municipal Environmental Research Laboratory. Cincinnati, Ohio 45268. EPA-600/2-78-114. July, 1978.

I. DIGESTER MIXING

1. Bauman, Peter G. and Huibregtse, Gregory L. "An Evaluation and Comparison of Digester Gas Mixing Systems." Fifty-third Conference of the Water Pollution Control Federation. 1980.
2. Dorr-Oliver Incorporated. Technical Reprints, Sanitary Engineering. Serial Nos. 1-999. Stanford, Connecticut.
3. Envirex Incorporated. PFT Water Quality Control Equipment, Binder No. 340. Waukesha, Wisconsin.
4. Gibbs, C.E., (ed.) Compressed Air and Gas Data, Ingersoll-Rand Company, New Jersey, 1971.
5. Holland and Chapman. Liquid Mixing and Processing in Stirred Tanks. Reinhold, New York. 1966.
6. Lemke, A.A. "Principles of Circulation Induced by an Unconfined Column of Gas Bubbles as Used in the CRP Accelerated Sludge digestion System." FMC Corporation. Environmental Equipment Division, July 1977.
7. Malina, J.F. "Thermal Effects on Completely Mixed Anaerobic Digestion." Water and Sewage Works. p. 52, January, 1964.
8. Monteith, H.D. and Stephenson, J.F., "Mixing Efficiencies in Full Scale Anaerobic Digesters by Tracer Methods," Wastewater Technology Centre, Environmental Protection Service, Department of the Environment, 1978.
9. Oldshue, J.Y., "Suspending Solids and Dispersing Gases in Mixing Vessel," Industrial and Engineering Chemistry, Volume 61, No. 9, 1969.
10. Parker, N.H., "Mixing," Chemical Engineering, June 8, 1964.
11. Smart, J., "An Assessment of the Mixing Performance of Several Anaerobic Digesters Using Tracer Response Techniques." Research Publication No. 72, Wastewater Treatment Section, Pollution Control Branch, Ontario Ministry of the Environment, 1978.
12. Torpey, W.N. and Melbinger N.R. "Reduction of Digester Sludge Volume by Controlled Recirculation." Journal of the Water Pollution Control Federation. Volume 39, p. 1464, 1967.
13. Uhl and Gray. Mixing Theory and Practice. Volume I and II. Academic Press. New York, New York. 1966; 1967.
14. Unknown Author. "Peripheral Mixing Turns Sludge Into Full Gas." The American City and County. July 1977.

I. DIGESTER MIXING (Continued)

15. Verhoff, F.H., Tenney, M.W. and Echelberger, W.F. "Mixing in Anaerobic Digestion." Biotechnology and Bioengineering. Volume XVI, p. 757, 1974.

J. THERMOPHILIC DIGESTION

1. Buhr, H.O. and Andrews, J.F. "Review Paper: The Thermophilic Anaerobic Digestion Process." Wter Research. Volume 11, p. 129, 1977.
2. Coulthard, T.L. and Townsley, P.M. "Thermophilic Processing of Municipal Waste." Paper No. 74.219, Canadian Society of Agricultural Engineers. 1974.
3. Fischer, A.J. and Greene, R.A. "Plant Scale Tests on Thermophilic Digestion." Sewage Works Journal. Volume 17, No. 7, 1945.
4. Garber, W.F. "Thermophilic Digestion at the Hyperion Treatment Plant." Journal of the Water Pollution Control Federation. No. 5, 1975.
5. O'Hara, George T. and Colbaugh, James E. "A Summary of Observations on Thermophilic Digester Operations." City of Los Angeles-Hyperion Treatment Plant. 1972-73. (Unpublished Report).

K. METHANE FERMENTATION

1. Barker, H.A. "Studies Upon the Methane Fermentation Process." Proceedings of the National Academy of Science. Volume 29, p. 184, 1943.
2. Heukelekian, H. and Heinemann, B. "Studies on Methane Producing Bacteria." Sewage Works Journal. Volume 11, p. 426, 471, 965. 1939.
3. Laurence, A.W. and McCarty, P.L. "Kinetics of Methane Fermentation in Anaerobic Treatment." Journal of the Water Pollution Control Federation. Volume 41, 1969.
4. Mylroie, R.L. and Hungate, R.E. "Experiments on Methane Bacteria In Sludge." Canadian Journal of Microbiology. Volume 1, p. 55, 1954.
5. McCarty, P.L. "The Methane Fermentation." Principles and Applications in Aquatic Microbiology. John Wiley and Sons, Inc., New York. 1964.
6. Vaseen, D.A. "New Method of Dual Media Fermentation Can Produce Quality Methane." Water and Wastes Engineering. Volume 14, No. 5, 1977.

L. DIGESTER GAS PRODUCTION AND UTILIZATION

1. Bauman, P.G. "Digester Methane Utilization Can Be Optimized." Water and Sewage Works. Volume 127, No. 11, 1980.
2. Brovko, N. and Chen, K.Y. "Optimizing Gas Production, Methane Content and Buffer Capacity in Digester Operation." Water and Sewage Works. p. 54, July, 1977.
3. Brown, Wade G. "Sludge Gas Utilization." Public Works. October, 1959.
4. Editor. "Gas Recovery Generates Income." The American City and County. July, 1977.
5. Editor. "Purification System Makes Pure Methane Out of Sewer Plant Gas." Pollution Engineering. Volume 11, No. 7, 1979.
6. Everett, A.D. "Digester Gas: Valuable Plant Fuel." Water and Sewage Works. May, 1975.
7. Fuhrman, R.E. "Sludge Gas Utilization." Sewage Works Journal. Volume 12, No. 11, 1940.
8. Glaub, J.C. and Diaz, L.F. "Biogas Purification Processes." (Cal Recovery Systems, Inc., Richmond, California). Biogas and Alcohol Fuels Production, Volume II. Journal of Waste Recycling. August, 1981.
9. Goepfner, J. and Hassilman, D.E. "Digester By-Product May Give Answer to Energy Problem." Water and Wastewater Engineering. No. 4, 1974.
10. Gould, R.H. "Present Status of Sludge Gas Utilization." Sewage Works Journal. Volume 19, No. 3, 1947.
11. Henrich, Anthony R. "Municipal Wastes to Vehicle Fuel." (Central Plants, Inc. Bio-Energy Division Commerce, California.) Biogas and Alcohol Fuels Production, Volume II. Journal of Waste Recycling. August, 1981.
12. Herpers, H. and Herpers, E. "Importance, Production and Utilization of Sewage Gases." KWC-Kohlenwasse-Stoffgase. Volume 72, p. 18, 1966.
13. Joseph, James. "Sludge Gas Becomes a Turbine Fuel." Water and Sewage Works. November, 1965.
14. Knopf, W.G. and Clarke, W.N. "Districts Minimize Energy Costs With Digester Gas." Journal of the Water Pollution Control Federation. Volume 51, No. 3, 1979.

L. DIGESTER GAS PRODUCTION AND UTILIZATION (Continued)

15. Martin, George. "Utilization of Sludge Gas in Moderate Sized Treatment Plants." Sewage Works Journal. Volume 14, No. 3, 1942.
16. Murphy, Cornelius, B. "Digester Off-Gas: An Untapped Energy Source." Forty-seventh Conference of the Water Pollution Control Federation. October, 1974.
17. Pruce, L.M. "Need for Greater Fuel Flexibility Spurs Gas-Turbine Developments." Power. Volume 124, No. 6, 1980.
18. Rimkus, R.R., Ryan, J.M. and Michuda, A.J. "Digester Gas Utilization Program." Journal of the Environmental Engineering Division. American Society of Civil Engineers. Volume 106, No. EE6, 1980.
19. Van Kleeck, L.W. "Sewage Sludge Gas-Its Collection and Use." Wastes Engineering. Volume 29, No. 2, 1958.
20. Ward, P.S. "Digester Gas Helps Meet Energy Needs." Journal of the Water Pollution Control Federation. Volume 46, No. 4, 1947.
21. Wesner, G. and Clarke, B. "There is A Lot of Energy in Digester Gas - Use It." The Bulletin. Volume 15, No. 1.

M. COGENERATION

1. Editor. "Cogeneration: Full Steam Ahead?" Power. 1978.
2. Strauss, S.D. "District Heating Links With Cogeneration." Power. Volume 123, No. 8, 1979.

N. POWDERED ACTIVATED CARBON

1. Hunsicker, M. "Powdered Activated Carbon Improves Anaerobic Digestion." Water and Sewage Works. Volume 123, No. 7, 1976.
2. Koch, C.M., Nelson, M.D. and DiMenna, R. "Wanted: More Gas, Less Slduge - Can PAC Help?" Water and Wastes Engineering. Volume 15, No. 6, 1978.
3. Spencer, R.R., Schuchrow, A.J. and Ferguson, J.F. "The Addition of Powdered Activated Cabon to Anaerobic Digesters: Effects on Methane Production." National Conference on Environmental Engineering Research, Development and Design. Seattle, Washington. July, 1976.
4. Ventetuolo, T. and Adams, A.D., "Improving Anaerobic Digester Operation with Powdered Activated Carbon." Water Pollution Control Federation, Deeds and Data. July, 1976.

O. CASE STUDIES

1. Battelle Pacific Northwest Laboratory, "Operational Characteristics of Anaerobic Digesters at Selected Municipal Wastewater Treatment Facilities in the United States" Prepared for the U.S. Department of Energy under Contract No. EY-76-C-06-1830, 1978.
2. Bergstedt, Dale C. "Energy Conservation and Recycling Program of the Metropolitan Sewer Board of the Twin Cities Area." Proceedings of the National Conference on Municipal Sludge Management. June, 1974.
3. Blodgett, J.H. "Accelerated Sludge Digestion Successful at Columbus, Ohio." Wastes Engineering. Volume 21, No. 12, 1950.
4. East Bay Municipal Utility District, Special District No. 1. Operating Report. July, 1979 - June, 1980.
5. East Bay Municipal Utility District. Fact Sheet. January 1, 1980.
6. Daly, W.J. "Cincinnati Plant Generates and Buys Power." Water and Sewage Works. Volume 125, No. 9, 1978.
7. Ghosh, S., Conrad, J.R., and Klass, D.L. "Anaerobic Acidogenesis of Wastewater Sludge." Journal of the Water Pollution Control Federation. Volume 47, p. 30, 1975.
8. Graef, Stephen P. "Anaerobic Digester Operation at the Metropolitan Sanitary Districts of Greater Chicago." Proceedings of the National Conference on Municipal Sludge Managements. June 1974.
9. Guarino, C.F. "Sludge Digestion Experiences in Philadelphia." Journal of the Water Pollution Control Federation. Volume 35, p. 626, 1963.
10. Haug, R.T. "Sludge Processing to Optimize Digestibility and Energy Production." LA/OMA. Journal of the Water Pollution Control Federation May, 1976.
11. Hunt, H. Harvey and Clarke, William N. "Design and Operation of the First Digester Gas Turbine in the USA." Journal of the Water Pollution Control Federation. Vol. 40, No. 7, 1968.
12. John Carollo Engineers. "Design Memorandum No. 11 - Digester Mixing," Prepared for County Sanitation Districts of Orange County, 1979.
13. John Carollo Engineers. "Digester Gas Utilization Study - Wastewater Solids Management Project." East Bay Municipal Utilities District. Interim Report. June, 1980.

14. John Carollo Engineers. Digester Gas Utilization Study Wastewater Solids Management Project. East Bay Municipal Utility District, Special District No. 1. April, 1981.
15. John Carollo Engineers. "Solids Handling Facilities For The 75 MGD Improved Treatment at Plant No. 2." County Sanitation Districts of Orange County. February, 1979.
16. McCarty, P.L., "Phase II Report of Laboratory Study on the Effects of Chemical Coagulants on Anaerobic Digestion of Richmond-Sunset Sludge." Prepared for CH₂M/Hill, 1974.
17. Mignone, Nicholas A. "Method of Reducing Sludge Produced from an Activated Sludge Plant." Environmental Technology Consultants, Inc., Springfield, VA. 1981. (Unpublished report).
18. Nelson, John, K. Supt. of Operations, Metropolitan Denver Sewage Disposal District No. 1. Denver, Colorado. 303-289-5941. Digester Operation Data, November, 1979 to October, 1980.
19. Nemhe, James. Madison Metropolitan Sewerage District, Madison, Wisc., Operation Data. July, 1978 to June, 1980.
20. O'Hara, G.T. and Colbaugh, J.E., "A Summary of Observations on Thermophilic Digester Operations." Proceedings of the 1975 National Conference on Municipal Sludge Management and Disposal. Information Transfer, Inc., Rockville, Maryland, 20852. June, 1974.
21. Smart, J. and Boyko, B.I. "Full Scale Studies on the Thermophilic Anaerobic Digestion Process." Research Report No. 59. Ontario Ministry of the Environment. Toronto, Ontario, Canada. 1977.
22. Spencer, R.R., Wong, A.L., Coates, J.A., and Ahlstrom, S.B. Operational Characteristics of Anaerobic Digesters at Selected Municipal Wastewater Treatment Facilities in the United States. U.S. Department of Energy Contract EY-76-C-06-1830. Prepared by Pacific Northwest Laboratory. Batelle Memorial Institute, Richland, WA. December, 1978.
23. Sykes, Robert M. Full Scale Field Demonstration of Unheated Anaerobic Contact Stabilization. U.S. Dept. of Energy Contract DE-AC02-80CS24310. Ohio State University. Columbus, Ohio.
24. Torpey, W.N. "Loading to Failure of a Pilot High-Rate Digester." Sewage Industrial Wastes. Vol. 27, p. 121, 1955.
25. Voshell, D. "Sludge Handling at Grand Rapids, Michigan Wastewater Treatment Plant." Journal of the Water Pollution Control Federation, Volume 38, p. 1506, (1966).

26. Water Utilities Dept., Dallas, Texas. "Specifications for the Purchase of Waste Digester Gas at the Central Wastewater Treatment Plant." Contract No. 75-34.
27. Wong, A.L., Rogers, E.A., English, C.J., Spencer, R.R. and Christensen, D.C. Production and Utilization of Methane from Anaerobic Sludge Digestion in the U.S. Wastewater Treatment Plant. U.S. Dept. of Energy Contract No. DY-76-C-06-1830. Prepared by Pacific Northwest Laboratory, Batelle Memorial Institute, Richland, WA. September, 1979.

P. MISCELLANEOUS

1. Anonymous "Upgrading Existing Wastewater Treatment Plants." EPA Process Design Manual. October, 1974.
2. Arora, M.L., Culp, G.L., Lineck, T.S. and Hinricks, D.J. "Retrofitting for Energy Savings." Water and Sewage Works. Volume 127, No. 6, 1980.
3. Bonner Jr., R.F. "Municipal Wastewater Treatment Plant Pioneers in Conservation Approach." American Society of Civil Engineers. Volume 80; No. 7, 1980.
4. Camp, T.R. and Stein, P.C. "Velocity Gradients and Internal Work in Fluid Motion." Journal of the Boston Society of Civil Engineers. p. 203, October, 1943.
5. Culp, Gordon, L. and Heim, Nancy Folks. Field Manual For Performance Evaluation and Troubleshooting at Municipal Wastewater Treatment Facilities. EPA-43019-78-001. Office of Water Program Washington, D.C. 20460. January, 1978.
6. Culp, Wesner and Culp. Final Value Engineering Report (Draft). City of Las Vegas, May, 1980.
7. Dames and Moore. Construction Costs For Municipal Wastewater Treatment Plants: 1973-1977. Water Pollution Control Engineering Services. Denver, Colorado. EPA 430/9-77-013, MCD-37. January, 1978.
8. Dick, R.I. and Ewing, B.B. "The Rheology of Activated Sludge" Journal of the Water Pollution Control Federation. Volume 39, p. 543, 1967.
9. Edison Electric Institute. Typical residential, commercial and industrial bills - Investor Owned Utilities. Winter 1981. Publication No. 04-81-06. Washington, D.C.
10. ENR Indexes. "Materials and Labor Cost Trends in the U.S." Engineering News Record. Volume 206, No. 12. March 19, 1981.
11. Fair, G.M., Geyer, U.C. and Okun, D.A. Water and Wastewater Engineering. Volume 2, John Wiley and Sons. New York, New York, 1968.
12. Finger, R.E. and Parrich, J. "Optimization of Grit Removal at a Wastewater Treatment Plant." Journal of the Water Pollution Control Federation. Volume 52, 1980.
13. Great Lakes Upper Mississippi River Board of State Sanitary Engineers. Recommended Standard For Sewage Works. 1978.

14. Hatsfield, W.D. "The Viscosity or Pseudo-Plastic Properties of Sewage Sludges." Sewage Works Journal. Volume 10, p. 3, 1938.
15. Jewell, W.J., et. al. "Bioconversion of Agricultural Wastes for Pollution Control and Energy Conservation." Prepared for the U.S. ERDA, Publication No. TID-27164, Cornell University, September, 1976.
16. John Carollo Engineers. "Preliminary Value Engineering Report. Wastewater Treatment Plant Expansion." City of Las Vegas. May, 1980.
17. Kerri, Kenneth, D. et. al. Operation of Wastewater Treatment Plants. A field study training program. Sacramento State College. EPA Technical Training Grant No. 5TT1-WP-16-03.
18. Lawrence, A.W. and McCarty, P.L. "A Unified Basis for Biological Treatment Design and Operation," Journal: Sanitary Engineering Division - ASCE. Volume 76, No. SA3, p. 757, 1970.
19. Metcalf and Eddy. Wastewater Engineering. McGraw-Hill Book Co. New York. 1972.
20. Perry, R.H. and Chilton, C.H., Chemical Engineers' Handbook. 5th Edition. McGraw-Hill Book Company. New York. 1973.
21. Sawyer, Clair N. and McCarty, Perry L. Chemistry for Sanitary Engineers. Second Edition. McGraw-Hill Book Company. New York. 1967.
22. USEPA. Construction Costs For Municipal Wastewater Treatment Plants. Office of Water Program Operations. Washington, D.C. 20460. MCD-37. January, 1978.
23. USEPA. Proces Design Manual For Upgrading Existing Wastewater Treatment Plants. Technology Transfer. Cincinnati, Ohio 45268. EPA 625/1-71-004a. October 1974.
24. Walker Process. Heat X. Digester Sludge Heating. Bulletin No. 25-S-82. Copyright 1979. Chicago Bridge and Iron Company.
25. Water Pollution Control Federation. Manual of Practice No. 11. Operation of Wastewater Treatment Plants. Washington, D.C. 20016. 1970.
26. Water Pollution Control Federation. Manual of Practice No. 8. Sewage Treatment Plant Design. Washington, D.C. 20016. 1972.

Q. INSULATION

1. Owens-Corning Fiberglass Corporation. Urethane Spray Systems General Application Guide. August, 1974. Publication No. 3-UF-6126-B.
2. Owens-Corning Fiberglass Corporation. Zip-Rib Tank Insulation System. May, 1978. Publication 1-1N-7610-B.

R. DIGESTER GAS SCRUBBING

1. Berry, Reginald I. "Membranes Separate Gas." Chemical Engineering. July 3, 1981.
2. Dingman, J.C., and Moore, T.F., "Compare DGA and MEA Sweetening Methods," Hydrocarbon Processing, Volume 47, No. 7, July, 1968.
3. Dingman, J.C., and Moore, T.F., "Gas Sweetening with Diglycolamine," Proceedings of the 1968 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
4. Ducksworth, G.L., and Geddes, J.H., "Natural Gas Desulfurized by the Iron Sponge Process," The Oil and Gas Journal, September 13, 1965.
5. Dunn, C.L., Freitas, E.R., Hill, E.S., and Sheeler, J.E.R., "Shell Reveals Commercial Data on Sulfinol Process," The Oil and Gas Journal, March 29, 1965.
6. Fails, J.C., and Harris, W.D., "Practical Way to Sweeten Natural Gas," The Oil and Gas Journal, July 11, 1960.
7. Goar, B.G., "Sulfinol Process Has Several Key Advantages," The Oil and Gas Journal, June 30, 1969.
8. Goar, B.G., "Today's Gas Treating Processes," Proceedings of the 1971 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
9. Goar, B.G., "Today's Sulfur Recovery Processes," Hydrocarbon Processing, Volume 47, No. 9, September, 1968.
10. Gustafson, D.J., and Healy, M.J., "Removal of Hydrogen Sulfide from Natural Gas/Carbon Dioxide Mixtures with Molecular Sieves," Proceedings of the 1968 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
11. Hegwer, A.M., and Harris, R.A., "Selexol Solves High H_2S/CO_2 Problems," Hydrocarbon Processing, Volume 49, No. 4, April, 1970.
12. Holder, H.L., "Diglycolamine - A Promising New Acid-Gas Remover," The Oil and Gas Journal, May 2, 1966.
13. Kohl, A.L. and Riesenfeld, F.C., Gas Purification, McGraw-Hill Book Company, Inc., New York City, New York, 1960.
14. Maddox, R.N., and Burns, M.D., "Designing a Hot Carbonate Process," The Oil and Gas Journal, November 13, 1967.
15. Maddox, R.N., and Burns, M.D., "Hot Carbonate - Another Possibility," The Oil and Gas Journal, October 9, 1967.

16. Maddox, R.N., and Burns, M.D., "Iron-Oxide Process Design Calculations," The Oil and Gas Journal, August 12, 1968.
17. Maddox, R.N., and Burns, M.D., "MEA Process to be Considered First," The Oil and Gas Journal, August 21, 1967.
18. Maddox, R.N., and Burns, M.D., "Physical Solution is the Key to These Treating Processes," The Oil and Gas Journal, January 8, 1968.
19. Mason, J.R., and Griffith, T.E., "A Case History - Conversion of a Sweetening Solution from MEA to DGA," Proceedings of the 1969 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
20. Nicklin, T., and Brunner, E. "How Stretford Process is Working," Hydrocarbon Processing, December, 1961.
21. Perry, C.R., "A New Look at Iron Sponge Treatment of Sour Gas," Proceedings of the 1970 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
22. Perry, C.R., "Basic Design and Cost Data on MEA Treatment Units," Portable Treaters, Inc., Odessa, Texas.
23. Perry, C.R., "Design and Operating Amine Units for Trouble Free Unattended Operation," Proceedings of the 1969 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
24. Rosen, W.D., "Here's a Quick Design Method for Amine Sweetening Plants," The Oil and Gas Journal, March 18, 1968.
25. Rushton, W.W., and Hays, W., "Selective Adsorption to Remove H₂S," The Oil and Gas Journal, September 18, 1961.
26. Scheirman, W.L., "Selection of Total Sulfur Recovery Systems Employing the Stretford Process," Proceedings of the 1972 Gas Conditioning Conference, University of Oklahoma, Norman, Oklahoma.
27. Shell, A.D., "Consider Basic Problems in Improving MEA Filtration," The Oil and Gas Journal, February 12, 1968.
28. Swaim, C.D., "Gas Sweetening Processes of the 1960's," Hydrocarbon Processing, Volume 49, No. 3, March, 1970.