An Assessment Tool for Managing Cost-Effective Energy Recovery from Anaerobically Digested Wastewater Solids
AN ASSESSMENT TOOL FOR MANAGING COST-EFFECTIVE ENERGY RECOVERY FROM ANAEROBICALLY DIGESTED WASTEWATER SOLIDS

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Abstract

Rapidly escalating energy cost over the past several years have resulted in renewed interest in the use of anaerobic digestion of wastewater solids for on-site energy recovery at municipal wastewater treatment facilities. Other factors may affect the decision to adopt an energy recovery system (ERS), however, including the need for pretreatment of the digester gas, the need to produce biosolids of a specified quality (e.g., Class A or B under the EPA 503 regulations), and the air emissions produced during the combustion of the digester gas. When these issues are combined with other factors such as the quantity and quality of the biogas produced onsite, the costs for purchased energy, as electricity or natural gas, and the cost of borrowing for the investment in the ERS, the decision to select one type of system over another can become daunting.

The number of commercially available ERS for energy recovery from digester gas is growing. The systems evaluated in this study included:

- Boilers
- Engine generators
- Turbines
- Microturbines
- Direct drive engines
- Stirling engines
- Fuel cells (phosphoric acid and molten carbonate)

The data acquired for use in this study were collected using detailed survey forms received from treatment plants, from site visits to selected facilities, from equipment suppliers and from the published technical literature.

A life-cycle assessment approach has been used in the development of a spreadsheet model (LCAMER) that enables users of the model to compare the relative economic merits of one ERS to another over the life of the systems. This resulting spreadsheet model from this study guides wastewater professionals through the technical and economic considerations for selecting the most effective ERS for their site-specific digester gas application. Using designated entry cells in the spreadsheet model, users can assess the relative effectiveness of different anaerobic digestion processes, gas pretreatment systems (GPSs), and ERs. The model outputs can be expressed as simple payback period or internal rates of return. Provision has been made to factor in the availability of monetary grants and, where applicable, the potential credits for reduction of greenhouse gas emissions. Emission factors for air pollutants have also been included in the spreadsheet model so that relative emission rates of common Criteria pollutants such as NOx, SOx and CO are estimated for any ERS investigated.

A detailed user’s manual accompanies the spreadsheet model, guiding the user through the cells in which data should or should not be entered was prepared. Several tutorial examples are provided to illustrate the use of the spreadsheet model.
The data and model testing suggest use of boilers and direct drive recovery units are likely to be among the most cost-effective ERSs, while fuel cell technology is not considered economically feasible under present economic circumstances.

**Benefits**

♦ Spreadsheet model enables wastewater treatment professionals to input site-specific plant operating data and local energy costs to assess the most cost-effective alternatives for recovering the energy in digester gas.

♦ Model can be used to determine the price for purchased energy at which a recovery system would become cost-effective based on the user’s acceptance criteria.

♦ Report provides a detailed review of the technical literature of various anaerobic digestion processes, digester GPSs and digester gas ERSs.

♦ Comprehensive literature database is a valuable reference source for subscribers interested in energy recovery from the anaerobic digestion process.

♦ Includes a listing of POTWs in North America that have ERSs, such as gas engine generators or fuel cells, and is a resource for subscribers to use when seeking first-hand information on operation of the ERSs.

**Keywords:** Digestion, biogas, methane, boilers, co-generation, microturbines, fuel cells, life-cycle assessment
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LIST OF ACRONYMS

ADS  Anaerobic Digestion System
AGAD Acid Gas phased Anaerobic Digestion
ATAD Autothermal Aerobic Digestion
BNR  Biological Nutrient Removal
BOD₅ Five-day Biochemical Oxygen Demand
COD  Chemical Oxygen Demand
ERS  Energy Recovery System
GPS  Gas Pretreatment System
LCA  Life Cycle Assessment
LCAMER Life Cycle Assessment Manager for Energy Recovery
LCC  Life cycle Cost
POTWs Public Owned Treatment Works
ppm  part per million
SRT  Solids Retention Time
TKN  Total Kjeldahl Nitrogen
TPAD Temperature Phased Anaerobic Digestion
TSS  Total Suspended Solids
VSR  Volatile Solids Reduction
WWTP  Wastewater Treatment Plant
EXECUTIVE SUMMARY

ES.1 Background

Anaerobic digestion of residual wastewater solids produces methane, which may be combusted on-site for energy production. Faced with increasing energy costs, wastewater treatment facilities are looking at opportunities for recovering the energy in the wastewater solids to help offset the need to purchase energy from off-site sources, whether as electricity or hydrocarbon fuels (e.g., natural gas). Efficient recovery of the energy in digester gas may have a significant impact on a treatment facility’s energy budget. Jewell (2005a), for example, has calculated that the energy recovered from cogeneration (combined heat and electricity production) might generate sufficient electricity to supply approximately 40% of the energy needed for aeration in a typical activated sludge process.

Much effort has been expended to optimize anaerobic digestion processes for methane production. A number of anaerobic sludge digestion technologies exist including low rate digestion, standard mesophilic high rate digestion (35°C, 95°F), two-stage high rate digestion (mesophilic), thermophilic digestion (55°C, 131°F), temperature-phased digestion (first stage at 55°C or 131°F, second stage at 35°C or 95°F) and metabolically-phased digestion (often called the acid-gas phase design). Each technology has advantages and disadvantages. For example, thermophilic digestion has the advantage of requiring shorter solids retention times and therefore smaller volume, but the disadvantages of being relatively more sensitive to upset, requiring more energy to heat, having a higher odor level, and producing solids that are more difficult to dewater. The temperature phased technology overcomes these issues, but at the cost of more capital investment.

Technologies for producing energy from biogas have also been examined but only recently have fuel cell and microturbine technologies developed sufficiently for practical consideration. Stationary fuel cells located on-site can now produce electricity directly, with higher efficiency and reduced emissions compared with conventional combustion process options. Each of the energy recovery technologies also has benefits and disadvantages. Boilers for example, are efficient thermally and relatively inexpensive compared to other energy recovery processes, but they do not produce electricity and they discharge pollutants such as carbon dioxide (CO₂) and nitrogen oxides (NOₓ) that may be considered undesirable. Engine generators produce both electricity and recoverable heat, but also emit criteria pollutants such as NOₓ and SOₓ. Microturbines and fuel cells emit low levels of pollutants, but their capital costs are expensive.

Wastewater treatment facilities face increasing energy costs and potential emissions regulation. There is a great need for comprehensive information on the most attractive (cost-effective) energy recovery options for these facilities, which can vary substantially depending on plant specific factors. The question – which energy recovery technology should be used under a given set of conditions? – cannot be readily answered, however, because there currently exists no method for appropriately comparing the necessary technical and economic factors.

Historically, wastewater treatment research has been narrowly focused on a particular stage of the treatment process (e.g., activated sludge or sludge digestion), a single media (e.g., air
emissions), or on a specific discharge (e.g., treated effluent) to the environment. However, systems level approaches, that go beyond localized analyses and draw a larger boundary to include the treatment plant as well as the environment and economy in which it operates, are necessary to evaluate the economic and sustainability implications of wastewater treatment operation options and to avoid making sub-optimal decisions (e.g. cross-media transfer of impacts, shifts of impacts from one stage to another).

Life Cycle Assessment (LCA) and Life Cycle Costing (LCC) are valuable systems level approaches. LCC looks at costs over the lifetime of a project. The traditional focus of LCC has been on private costs, those that are incurred solely by the project owner. However, costs external to the project owner but internal to society as a whole (externalities) can be incorporated. This research developed a spreadsheet-based model called LCAMER (Life Cycle Assessment Manager for Energy Recovery) that provides this information, facilitating a comparison of energy recovery alternatives, such as co-generation, fuel cells and microturbines. This modeling tool identifies the economically most attractive alternatives for a wide range of factors such as energy costs, regulatory conditions, wastewater plant capacity and operations, and social values.

The objectives of this project were to:

1. Develop a model for comparing the benefits and costs of alternatives for energy recovery throughout the lifetime of the treatment facility as a function of plant factors and external parameters.

2. Use the model to develop cost-effectiveness response surfaces for existing and novel sludge treatment and energy recovery alternatives.

3. Predict the impact on overall plant operation due to changes in solids treatment and energy recovery technologies.

4. Predict the most economically viable sludge treatment and energy recovery alternatives for given input sets.

5. Provide WERF subscribers a copy of the model with a user’s manual for use at their facilities.

**ES.2 Methodology**

**Literature Search**

Sources of information and data for this study were obtained from computerized literature searchers and from detailed questionnaires sent to municipalities identified as having either an anaerobic digestion process or digester gas energy recovery process of interest, or both.

**Treatment Facility Data Collection**

An initial detailed data collection template was sent out in early June 2003 to 15 target wastewater treatment facilities. The template was designed to cover many aspects of this process, in terms of the type of treatment processes used, details about digester design (e.g., insulation types) and operations (e.g., retention time, volatile solids reductions, gas production and quality), energy use and recovery data, distribution of energy use on-site using data such as numbers and...
sizes of pump motors, and capital and operating cost data for the anaerobic digestion and recovery system capital an O&M costs. The survey was detailed, but such data are required to develop the process and costing functions used in the life-cycle assessments, and to develop the software that enables Public Own Treatment Works (POTWs) to determine the optimum energy recovery strategy. A second survey was sent out in October 2003 to expand the database used for developing values used in the non-monetary and monetary functions in the cost-effectiveness model. The second survey was slightly modified from the first to eliminate redundancies, and to include data such as whether or not the facility was located in a designated ozone non-attainment area, which could influence the choice of energy recovery process. A database was created in MS-Access to allow for data querying for different combinations of search terms and constraints.

**Non-Monetary Digestion and Energy Recovery Process Functions**

For LCAMER model development, relationships between many diverse input variables must be defined mathematically. Collectively, the input variables can be used in either non-monetary functions, relating digester gas production and energy value, as well as energy recovery terms. The cost functions make use of concepts such as present and future worth, capital and operating costs and revenue streams.

The non-monetary (process) functions have four variable types: input, operational, intermediate, and output. *Input variables* are those that are fully independent of the given unit process (e.g. anaerobic digestion). *Operational variables* are those that may be independently controlled, but are not independent of the unit process; they are very much an integral part of that process. They are site-specific and can vary from one facility to the next. *Intermediate variables* are dependent to input and/or operational variables, and independent to output variables. *Output variables* are dependent upon the input, operational, and/or intermediate variables within the same unit process. It should be noted that output variables for one process may be input variables for another process further downstream. Table ES-1 is an example of the non-monetary variables involved with the energy recovery processes using digester gas.

<table>
<thead>
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<th>Operating variables</th>
<th>Intermediate variables</th>
<th>Output variables</th>
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<td>Total gas flow rate, gpd (m³/d)</td>
<td>Thermal energy, Btu/d (MJ/d)</td>
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<td>% Methane</td>
<td>Operating efficiency</td>
<td>Gas energy content Btu/scf (MJ/m³)</td>
<td>Electrical energy, Btu/d (MJ/d)</td>
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<td>Energy Recovery process</td>
<td></td>
<td>% Thermal Energy Recovery efficiency</td>
<td>Mechanical energy, Btu/d (MJ/d)</td>
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<tr>
<td></td>
<td></td>
<td>% Electrical Energy Recovery efficiency</td>
<td>Non-useful energy, Btu/d (MJ/d)</td>
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<tr>
<td></td>
<td></td>
<td>% Mechanical Energy Recovery efficiency</td>
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**Table ES-1. Technical Energy Recovery Variables.**

Monetary Functions

The generalized life cycle cost is given in Equation ES-1.

\[
LCC = CC + OC - R - S
\]

*(ES-1)*

Where: \(\text{LCC} = \text{net present life cycle cost,}\)
CC = net present capital cost,
OC = net present operating cost,
R = net present revenue,
S = net present savings,
s = scenario being investigated.

Capital costs are almost always incurred prior to the beginning of plant operation. The purchase of replacement equipment and/or technologies may occur at any point within the time horizon being investigated. Annual interest rates must be considered in the calculations, as well as inflation, if so desired. Operating costs are incurred on a regular basis, whether daily, weekly, monthly, or some other time period. Fixed operating costs are indexed in $ terms and do not change in relation to a given parameter. Variable operating costs are divided into four sub-cost types consisting of labor, chemicals, maintenance and energy. Revenues are divided into annual revenues, which may come from off-site sale of heat or electricity, and from salvage value. In the future, greenhouse gas emission reduction credits may be obtained for reducing the need for upstream generation of electricity, but these are not in place now. Savings may be even more important than revenues for many facilities and arise from reduced expenditures for natural gas and electricity when biogas is used for heat and electricity production.

Several unique considerations existed when applying the LCA approach to anaerobic digestion and energy recovery. These included:

♦ Development of life-cycle costing functions for anaerobic sludge digestion and energy recovery at POTWs
♦ Development of a LCA-based analytical model for anaerobic sludge digestion and energy recovery
♦ Development of LC monetization functions for calculating external costs
♦ Application of an environment- and economy-wide approach for treatment plants.

Effect of Supernatant Quality from Innovative Digestion Process Upgrade on Liquid Treatment

The effect of anaerobic digester supernatant on the operation of liquid treatment processes has been recognized for some time. What is less clear is the effect of changing from one anaerobic technology, such as high-rate mesophilic digestion to one of the more innovative anaerobic technologies. The effect of upgrading technologies was evaluated using a combination of literature review and spreadsheet estimations.

ES.3 Results

Review of Anaerobic Digestion, Gas Pretreatment and Energy Recovery Processes

Anaerobic digestion, digester gas pretreatment and energy recovery processes identified in the literature searches are summarized in Table ES-2. Digestion processes listed in Table ES-2
other than single- and two-stage mesophilic digestion are considered as innovative anaerobic digestion processes.

**Survey Results**

Mesophilic anaerobic processes are the predominant form of anaerobic wastewater solids digestion. Digestion processes listed in Table ES-3 other than single- and two-stage mesophilic digestion are considered as innovative anaerobic digestion processes. While the detailed survey forms were sent to other facilities, particularly the ones with processes deemed innovative, responses were not forthcoming in the available time period.

The distribution of responding treatment plants with different types of energy recovery systems (ERS) is provided in Table ES-4. Boilers were the most common form of energy recovery from biogas, followed by engine-generators. These technologies represent the more traditional digester gas energy recovery processes. Microturbines and direct drive units appear to have growing acceptance as energy recovery technologies.
Table ES-2. Possible Component Options Identified for Energy Recovery from Anaerobically Digested Wastewater Solids.

<table>
<thead>
<tr>
<th>Anaerobic Digestion</th>
<th>Gas Pretreatment</th>
<th>Energy Recovery</th>
</tr>
</thead>
<tbody>
<tr>
<td>Single-stage mesophilic, 95°F (35°C)</td>
<td>no pretreatment</td>
<td>biogas flaring (no energy recovery)</td>
</tr>
<tr>
<td>Two-stage¹ mesophilic</td>
<td>Moisture removal only</td>
<td>combustion in boiler with thermal energy recovery (steam and/or hot water)</td>
</tr>
<tr>
<td>Single-stage thermophilic, 131°F (55°C)</td>
<td>moisture and H₂S removal⁶</td>
<td>combustion in boiler with thermal energy recovery (hot water) and use of superheated steam in steam turbine/generator for electricity generation</td>
</tr>
<tr>
<td>Two-stage thermophilic</td>
<td>Moisture, H₂S, and particulate removal</td>
<td>combustion in engine generator for electricity generation and thermal energy recovery (exhaust gas and/or hot jacket water)</td>
</tr>
<tr>
<td>Two-phase² mesophilic</td>
<td>Moisture, H₂S, CO₂, and particulate removal</td>
<td>combustion in direct drive engine for mechanical work and thermal energy recovery</td>
</tr>
<tr>
<td>Two-phase thermophilic</td>
<td></td>
<td>combustion in transportation system engine (not investigated as option in report)</td>
</tr>
<tr>
<td>Temperature phased³ (meso-thermo)</td>
<td></td>
<td>combustion in turbine for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>Temperature phased (thermo-meso)</td>
<td></td>
<td>combustion in microturbine (&lt;150 kW) for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>Multi-stage⁴ mesophilic</td>
<td></td>
<td>combustion in Stirling engine for mechanical work and thermal energy recovery</td>
</tr>
<tr>
<td>Multi-stage thermophilic</td>
<td></td>
<td>phosphoric acid stationary fuel cell⁷ for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>multi-stage temperature phased⁵</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

¹ “two-stage” denotes the use of two digesters in series; not the separation of fermentation and methanogenesis
² “two-phase” denotes the physical separation of fermentation and methanogenesis into separate digesters, such that the growth conditions for each (pH, etc.) is optimized
³ two digesters in series, each with a different operating temperature range
⁴ more than two digesters in series
⁵ more than two digesters in series; not all having the same operating temperature range (e.g. meso-thermo-meso)
⁶ the pressure required to drive the gas pretreatment process is often specific to the patented technology; as such, compression shall not be considered to be a separate factor for gas pretreatment at this time (i.e. if the process requires compressing the gas to 200 psi, it must be done)
⁷ the PAFC is the only commercially available fuel cell; other fuel cell types that may be used for stationary purposes include the proton exchange fuel cell, the molten carbonate fuel cell, and the solid oxide fuel cell

Table ES-3. Types of Digestion Systems in Survey Responses.

<table>
<thead>
<tr>
<th>Type of Anaerobic Digestion System</th>
<th>Number in Survey</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mesophilic – High Rate Single Stage</td>
<td>13</td>
</tr>
<tr>
<td>Mesophilic – High Rate Two Stage (mixed/unmixed secondary digesters)</td>
<td>11/13</td>
</tr>
<tr>
<td>Thermophilic – High Rate Single Stage</td>
<td>2</td>
</tr>
<tr>
<td>Thermophilic – High Rate Multi Stage</td>
<td>1</td>
</tr>
<tr>
<td>Temperature phased (M/T or T/M)</td>
<td>2</td>
</tr>
<tr>
<td>Acid-Gas phased</td>
<td>0</td>
</tr>
<tr>
<td>TOTAL</td>
<td>42</td>
</tr>
</tbody>
</table>

M = Mesophilic, T = Thermophilic

<table>
<thead>
<tr>
<th>Type of Energy Recovery System</th>
<th>Number in Survey</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boilers</td>
<td>34</td>
</tr>
<tr>
<td>Engine-generators</td>
<td>16</td>
</tr>
<tr>
<td>Microturbines</td>
<td>6</td>
</tr>
<tr>
<td>Fuel Cells (Molten carbonate or phosphoric acid)</td>
<td>2 + 2 a</td>
</tr>
<tr>
<td>Direct Drive</td>
<td>7</td>
</tr>
<tr>
<td>Other</td>
<td>6 b</td>
</tr>
<tr>
<td><strong>TOTAL</strong></td>
<td><strong>71 c</strong></td>
</tr>
</tbody>
</table>

a 2 plants surveyed use biogas in fuel cells. Additionally, Terminal Island uses natural gas only in fuel cells, King County in Seattle WA has fuel cells (plant was visited by did not complete a survey)
b Includes 1 flare only; 1 sending biogas through pipeline; 1 steam turbine; 1 gas turbine, 1 combined cycle turbine and 1 pelletizing burner
c Total number of systems is greater than the number of plants surveyed because some plants have more than one type of energy recovery system.

**Efficiency of Energy Recovery Systems**

Pertinent information including electrical and thermal energy recovery efficiencies, \( \text{H}_2\text{S} \) contamination limits, and the minimum biogas production required (based on the smallest available commercial size) for the energy recovery technologies are summarized in Table ES-5. The numbers presented were conservatively chosen from the lower end of ranges of common values in literature, manufacturer and survey data. Boilers are still one of the most cost-effective methods for recovering energy, having no gas pretreatment requirements, high thermal recovery efficiency and low capital and operating costs. The drawback to boilers is that they recover thermal energy only, and so may not be as useful in the warmer temperate or tropical climates. Direct drive units are more energy efficient than others that first convert the chemical energy of biogas to electricity, and then again to mechanical energy.

Table ES-5. Energy Recovery Efficiencies and \( \text{H}_2\text{S} \) limits of Technologies.

<table>
<thead>
<tr>
<th>Technology (ERS)</th>
<th>Electrical Efficiency (%)</th>
<th>Recoverable Thermal Efficiency (%)</th>
<th>Smallest size (MW)</th>
<th>Minimum Biogas Required scf/d (m³/day)a</th>
<th>( \text{H}_2\text{S} ) contaminant limit (ppmv)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler</td>
<td>-</td>
<td>80</td>
<td>0.3b</td>
<td>52,260 (1480)</td>
<td>NSd</td>
</tr>
<tr>
<td>Gas Engine Electricity/ Cogeneration</td>
<td>33</td>
<td>40</td>
<td>0.5</td>
<td>215,200 (6094)</td>
<td>100</td>
</tr>
<tr>
<td>Gas Engines Direct Drive/ Heat Recovery</td>
<td>35c</td>
<td>40</td>
<td>0.5</td>
<td>202,900 (5746)</td>
<td>100</td>
</tr>
<tr>
<td>Microturbin</td>
<td>27</td>
<td>40</td>
<td>0.03</td>
<td>15,790 (447)</td>
<td>70 000 (NS)d,e</td>
</tr>
<tr>
<td>Gas turbine</td>
<td>30</td>
<td>40</td>
<td>0.5</td>
<td>236,700 (6704)</td>
<td>25</td>
</tr>
<tr>
<td>Steam turbine</td>
<td>25</td>
<td>20</td>
<td>0.1</td>
<td>56,820 (1609)</td>
<td>500</td>
</tr>
<tr>
<td>Combined cycle</td>
<td>50</td>
<td>40</td>
<td>1</td>
<td>284,100 (8045)</td>
<td>25</td>
</tr>
<tr>
<td>Phosphoric acid fuel cell</td>
<td>35</td>
<td>45</td>
<td>0.2</td>
<td>81,150 (2298)</td>
<td>200f</td>
</tr>
<tr>
<td>Molten carbonate fuel cell</td>
<td>50</td>
<td>30</td>
<td>1</td>
<td>284,100 (8045)</td>
<td>1</td>
</tr>
<tr>
<td>Stirling Engines</td>
<td>25</td>
<td>45</td>
<td>0.055</td>
<td>31,250 (885)</td>
<td>NSd</td>
</tr>
</tbody>
</table>

a based on smallest technology size, 60% methane in biogas, and a 961.41 Btu/scf (35.8MJ/m³) methane energy content
b based on a 30bhp sized boiler
c mechanical conversion efficiency for direct drives is 35%
d NS = not sensitive to the \( \text{H}_2\text{S} \) concentrations expected in biogas
e biogas drying and compression are required
f with the gas processing unit supplied for use with digester gas
Development of LCAMER Spreadsheet Model

The LCAMER model for the biogas production rate included as part of this project requires an estimate of the volatile solids reduction efficiency. A semi-empirical steady state model was developed that would accurately predict volatile solids (VS) Reduction in anaerobic digesters as a function of less complex and more readily available inputs, including temperature, sludge retention time (SRT), fraction of each sludge type (primary sludge and waste activated sludge), pH and VS load. The estimated VS Reduction then would be used to determine the biogas production rate in the digester.

The VSR model developed for this project is described in detail by Kalogo et al. (2005), and is not reproduced here. In essence it makes use of the parameters noted above, as well as hydrolysis constants, to estimate the VS Reduction. VS Reduction values predicted by the semi-empirical model of Kalogo et al. (2005) were compared to VS Reduction values calculated by 34 wastewater Treatment plants (WWTPs), either reported in surveys or the literature, representing seven different anaerobic technologies. The results are provided in Figure ES-1. The model predicted well the performance of the plants with the difference between the model responses and the plants performances varying from -1% to +8%. In addition, a limited number of comparisons of the International Water Association’s Anaerobic Digestion Model No. 1 (ADM1) (Batstone et al., 2002) are depicted in Figure ES-1. The data indicated that the less complex, semi-empirical model predicted the VS Reduction in the digestion process at least as well as ADM1 over the range of conditions tested. The semi-empirical model, being less complex than ADM1, thus offers process engineers and operators a simpler alternative for predicting VS Reduction in anaerobic digesters.

Figure ES-1. Predictive Capability of Semi-Empirical VSR Model for Estimation of VS Reduction.

The entire LCAMER spreadsheet tool consists of seven worksheets. Two of the worksheets (labeled Technical and Economics) are those to which a model user would enter plant site-specific data. Of the remaining worksheets, one briefly describes the contents of each worksheet, while the remaining four worksheets contain calculations and data that are used by the main Technical and
Economics worksheets. A detailed user manual has been prepared for use with the spreadsheet model. The model is available in both U.S. and international metric unit versions.

**LCAMER Model Application**

A model plant consisting of the assumptions and variables described for the medium sized plant in the previous section was examined for energy recovery potential. Electricity and heat production using all ERS were calculated based on the efficiencies shown above in Table ES-5, and the biogas production rate from the LCAMER model. The results are shown in Figure ES-2, along with the plant electricity and digester heating demands. ERS were sized using commercially available sizes and all biogas produced was assumed to be used in the ERS (volume loss through gas treatment was assumed to be negligible). Therefore, in some cases because of fixed available sizes, extra energy recovery capacity was available which was met by supplementing with natural gas. The results below, however, show only the energy production from the biogas.

![Figure ES-2. Heat and Electricity Production for Various Energy Recovery Systems.](image)

With the selected input parameters in this scenario, all the cogeneration technologies were predicted to produce between 35-70% of the plant’s electricity demand. All technologies should produce enough heat to meet the summer demand except for steam turbines and the molten carbonate fuel cells, due to their lower thermal efficiencies. Based on the assumptions used in the model, none of the cogeneration technologies produce enough heat to meet the winter heat demand for digesters. In this scenario, if cogeneration technologies are used, boilers fuelled with natural gas...
will likely be needed to supplement winter heating requirements. These results were generated using temperature data for a plant in Wisconsin. Plants located further south will likely require less winter heating while those in Canada may require more winter heating.

With respect to the economic viability of energy recovery, a sensitivity analysis revealed that nearly half the uncertainty associated with the economic viability was related to the price of purchased electricity, while the unit annual O&M cost for the gas pretreatment system accounted for over one-third of the economic uncertainty.

Jewell (2005b) has calculated, however, that the assumptions used in heat efficiency may reflect inadequate digestion tank insulation rather than a true lack of sufficient energy to maintain the digestion process. The calculations indicate, based on the energy recovered from the raw wastewater solids, and energy recovered both from cogeneration and by heat exchangers, that a feed solids concentration of 3% volatile solids should have a heating potential of approximately 45°C when incorporating heat exchangers, cogeneration and a suitably insulated tank. Use of a more efficient insulation system in the spreadsheet model would result in the heat output energy in the biogas matching more closely the heat demand, especially in the colder winter months.

Potential Applications of the LCAMER Spreadsheet Model
The LCAMER spreadsheet model can be a useful tool for helping to evaluate potential adoption of biogas energy recovery processes. A number of potential applications are indicated below:

♦ Evaluation of the benefit of adding a process such as sonification or MicroSludge™ to improve volatile solids reduction, and hence biogas production;

♦ Feasibility of switching from boilers to another form of combined thermal and mechanical/electrical energy for onsite use;

♦ Feasibility of switching from mesophilic to thermophilic in a plant expansion, with added energy recovery processes;

♦ Feasibility of adopting an energy recovery process considering that purchased energy prices for electricity and natural gas will rise over a number of years

Results of Life Cycle Assessments for Digestion, Gas Pretreatment and Energy Recovery Options
Relative valuations of the different processes were accomplished by comparing the net present valuations of the options to that of a baseline scenario. For example, the baseline condition assumed for one assessment was: concrete conventional digesters with a one-stage mesophilic anaerobic digestion process, and boilers for combusting digester gas to heat sludge, with flaring of the excess digester gas. Alternative scenarios were composed of concrete conventional digesters, with either a mesophilic one-stage digestion process or a one-stage thermophilic process, an activated carbon gas pretreatment system, and one alternative energy recovery. Boilers, Steam turbines, and Stirling engines were considered not to require gas pretreatment. When the thermal energy recovered from those energy recovery options was not sufficient for heating sludge, boilers were applied to combust digester gas for supplementary heat.
The LCC analysis results, in terms of net present value, are presented in Figure ES-3. The differences in alternative LCCs from the baseline LCC (i.e., Delta LCC) are lower than the baseline for all alternatives except the PA fuel cell and MC fuel cell options. Operation of the anaerobic digestion process at thermophilic temperatures improved the economics for all the scenarios tested, including the fuel cells. The Delta LCC is more favorable for thermophilic than mesophilic digestion, in spite of higher energy consumption, due to reduced capital costs as a result of the shorter retention time needed for thermophilic compared to mesophilic digestion. Based on the calculated negative values for Delta LCC, most combined heat and power technologies should be economically viable. Due to their high capital costs, fuel cells are not yet economically feasible for WWTPs without capital grants.

**Figure ES-3. Comparisons of Delta Life Cycle Costs (in Net Present Values in 2004 $US) for Selected Alternative Scenarios.**
INTRODUCTION

1.1 Rationale

Municipal wastewaters must be treated to protect human health and receiving waters while, increasingly, providing water for reuse. Treatment processes require energy and produce waste solids (Figure 1-1). Additional solids treatment is needed to produce biosolids which may be beneficially land applied. Processes exist for converting waste solids into biosolids; some require energy, others produce energy. Those processes that produce energy are intuitively appealing. The produced energy can be applied to the wastewater treatment energy requirements and, under appropriate conditions, reduce cost. Assuming that a suitable solids treatment train can be constructed to achieve biosolids stabilization criteria, the remaining questions are how to maximize energy production from the solids treatment process and minimize overall treatment costs.

![Figure 1-1. Abbreviated Municipal Wastewater Treatment Schematic.](image)

The answers to these questions have been pursued for many years. Anaerobic digestion of residual wastewater solids produces methane, which may be combusted on-site for energy production. Faced with increasing energy costs, wastewater treatment facilities are looking at opportunities for recovering the energy in the wastewater solids to help offset the need to purchase energy from off-site sources, whether as electricity or hydrocarbon fuels (e.g., natural gas). Efficient recovery of the energy in digester gas may have a significant impact on a treatment facility’s energy budget. Jewell (2005a), for example, has calculated that the energy recovered from cogeneration (combined heat and electricity production) might generate sufficient electricity...
to supply approximately 40% of the energy needed for aeration in a typical activated sludge process.

Much effort has been expended to optimize anaerobic digestion processes for methane production. Technologies for producing energy from biogas have also been examined but only recently have fuel cell and microturbine technologies developed sufficiently for practical consideration. Stationary fuel cells located on-site can now produce electricity directly, with higher efficiency and reduced emissions compared with conventional combustion process options. Wastewater treatment facilities face increasing energy costs and potential emissions regulation. There is a great need for comprehensive information on the most attractive (cost-effective) energy recovery options for these facilities. The research completed in this project has created a spreadsheet model that provides this information, facilitating a comparison of energy recovery alternatives, such as but not limited to, co-generation, fuel cells and microturbines. The model identifies the economically most attractive alternatives for a wide range of factors such as energy costs, regulatory conditions, wastewater plant capacity and operations. The literature review that follows amplifies the interactions of the various factors addressed in this research.

1.2 State of the Art

1.2.1 Wastewater Solids Treatment Technologies

Wastewater solids treatment is required to reduce volume and stabilize the solids. If the treated solids are to be land applied, they must meet the appropriate regulatory requirements (see for example U.S. EPA, 1996; Ontario MOE, 1998). These requirements specify the maximum concentrations of constituents such as metals and pathogens that may be contained in the solids. In some cases (U.S. EPA, 1996), the physical conditions and/or technologies that may be used to achieve pathogen reduction are also specified.

The biological processes that may be used for sludge treatment are classified as either aerobic (requiring oxygen) or anaerobic (occurring in the absence of oxygen). Several aerobic processes are available that achieve target volatile solids destruction and pathogen reduction. Autothermal thermophilic aerobic digestion (ATAD) processes can maintain thermophilic temperatures (55°C, 131°F) without external heat but not one aerobic digestion process, ATAD included, allows for energy recovery from the waste organic material. Instead, additional energy must be expended to provide aeration, which, in essence, wastes the energy content of the solids by converting the organic materials to carbon dioxide (CO₂). Post-treatment of anaerobically digested solids by ATAD has some potential as a treatment scheme (Jewell, 2005c).

If energy recovery from solids digestion is desired, anaerobic treatment processes must be used. A consortium of anaerobic microorganisms hydrolyzes the solids, ferments the more complex molecules to simpler organic acids, and ultimately converts the resulting acetic acid to methane (CH₄) and CO₂. Hydrogen (H₂) produced by the fermentative microorganisms is also consumed in conjunction with CO₂ to produce CH₄. The challenge of any anaerobic system is to balance the rate of organic acid production by the fermentative organisms with the rate of acetic acid and H₂ consumption by the methanogenic organisms. If acid formation is too rapid, pH may drop, inhibiting methanogenic activity and possibly causing performance failure. Proper design and operation with particular attention to alkalinity requirements are required to minimize the potential for upsets.
A number of anaerobic sludge digestion technologies exist including low rate digestion, standard mesophilic high rate digestion (35°C, 95°F), two-stage high rate digestion (mesophilic), thermophilic digestion (55°C, 131°F), temperature-phased digestion (first stage at 55°C or 131°F, second stage at 35°C or 95°F) and metabolically-phased digestion (often called the acid-gas phase design). Each technology has advantages and disadvantages. For example, thermophilic digestion has the advantage of requiring shorter solids retention times and therefore smaller volume, but the disadvantages of being relatively more sensitive to upset, requiring more energy to heat and producing solids that are more difficult to dewater. The temperature phased technology overcomes these issues, but at the cost of more capital investment.

Perhaps the biggest drawback of existing anaerobic sludge digestion technologies is the requirement to produce methane. The potential for microbial imbalance is thermodynamically pre-ordained. If methanogenic organisms cannot consume H₂ fast enough, fermentative organisms will use organic electron acceptors and form particularly difficult to degrade compounds such as propionic or butyric acids which will drop the system pH. The metabolically-phased design attempts to mitigate this problem by allowing fermentation processes to occur in one stage to rapidly produce organic acids which are then converted to methane in a separate second stage.

While the metabolically-phased design may lead to more stable operation, the requirement to produce methane has not been removed. The system remains thermodynamically limited because H₂ is not removed in the first phase. Computer simulations that incorporate thermodynamic limitations (using the model developed by Bagley and Brodkorb, 1999) predict that systems which remove H₂ during the fermentation phase will perform more robustly and produce more acetic acid than other systems. The development of anaerobic sludge digestion technology to produce H₂ is underway but until the technology is sufficiently mature, the state of the art is to produce methane using an appropriate methanogenic digester.

1.2.2 Energy Recovery

The methane-containing biogas produced from anaerobic solids digestion is potential energy. A number of alternatives have been examined and implemented to recover the potential energy in the biogas. Until recently, these alternatives were combustion based. The simplest alternative is to combust the biogas in a boiler, and use the steam for heating, in particular the digesters themselves. A fraction of the cost of energy required to heat the digesters could be recovered, however overall cost-effectiveness depended on the capital, operational and maintenance (O&M) costs of the gas collection and combustion facilities as well.

Another alternative that has been implemented is cogeneration to produce on-site electricity as well as heat. A cheaper version of this alternative, from a capital cost standpoint, is to send the biogas off-site to a nearby utility which can then use less virgin natural gas to produce electricity. The municipal WWTP may then receive a discounted rate on their electricity. Perhaps the largest example of this approach is the Hyperion plant in Los Angeles which sends 16 MW worth of biogas to the utility and then receives a discounted rate on 16 MW of the 20 MW required to operate the facility.

A recent energy recovery alternative is the use of on-site fuel cells to produce electricity without combustion. At least five municipal WWTPs in the U.S. (Columbia Boulevard plant in Portland, OR; Terminal Island plant in Los Angeles, CA; Yonkers, NY; Las Virgenes Municipal
Water District (Calabasas, CA); King County, WA, as well as others in Europe and Japan now use 200 kW fuel cells on-site. The biogas is scrubbed and reformed to hydrogen for use in the fuel cell. Although the total amount of CO₂ produced from a fuel cell is the same, per cubic meter (m³) of biogas used, as that from combustion, reduced air emissions are generated due to the lower temperature operation. Furthermore, the energy recovery is, in theory, higher, so the CO₂ emissions per energy unit are less.

The state of the art for energy recovery, then, varies substantially depending on plant specific factors. The question – which energy recovery technology should be used under a given set of conditions? – cannot be readily answered, however, because there currently exists no model for appropriately comparing the necessary technical and economic factors.

1.2.3 Systems Level Analyses

While significant ground-breaking research on wastewater treatment has occurred in the past, there have been few studies that integrate the results into systems level analyses. Historically, wastewater treatment research has been narrowly focused on a particular stage of the treatment process (e.g., activated sludge or sludge digestion), a single media (e.g., air emissions), or on a specific discharge (e.g., treated effluent) to the environment. Narrowly focused studies are useful to answer specific questions and provide data for broader studies. However, systems level approaches, that go beyond localized analyses and draw a larger boundary to include the treatment plant as well as the environment and economy in which it operates, are necessary to evaluate the economic and sustainability implications of wastewater treatment operation options and to avoid making sub-optimal decisions (e.g. cross-media transfer of impacts, shifts of impacts from one stage to another). Ausubel (1998) and NAE (1999) provide both introductory and detailed information on the current status of research in the emerging field of systems level analysis.

LCA and Life Cycle Costing (LCC) are valuable systems level approaches. The LCA literature, however, has focused primarily on process models of commercial products, capturing the environmental effects of producing, transporting, using, and disposing of the product during its lifetime (Curran, 1996). Included in this series of activities are the effects of extracting the raw materials, producing the energy and other resource inputs, etc. Within North America, the Society for Environmental Toxicology and Chemistry (SETAC) and the U.S. Environmental Protection Agency (EPA) have played the leading roles in standardizing the ‘conventional’ LCA method (SETAC, 1993, Vigon et al., 1993).

LCC looks at costs over the lifetime of a project. LCC was developed by the U.S. Department of Defence during the 1960s for procurement situations so as to consider lifetime costs of products and to avoid situations wherein costs were transferred from one stage of the product life cycle to another (e.g. lower construction cost but increased maintenance costs). The traditional focus of LCC has been on private costs, those that are incurred solely by the project owner. However, costs external to the project owner but internal to society as a whole (externalities) can be incorporated.

One method for LCA is the Economic Input-Output Life Cycle Assessment (EIO-LCA) model developed at Carnegie Mellon University. The model is based on a 485 by 485 sector Input-Output (I-O) technical coefficient matrix of the U.S. economy that is augmented with an environmental matrix (Hendrickson et al., 1998). Economic Input-Output (I-O) analysis has a long
history and was originally formalized by Wassily Leontief (Leontief, 1966) for which he won a Nobel Prize in 1975. However, the EIO-LCA model is the first application of the method to LCA. The model has the capacity to evaluate all inter-industry or inter-commodity transactions in the economy. The model output includes the economic and environmental implications throughout the entire economy resulting from a demand for the product or service produced by any sector (Hendrickson et al., 1998).

The implications of LCA include direct implications (resulting from the industry itself) and indirect (related to its entire supply chain). The EIO-LCA model is quick, comprehensive, and available free of charge on the Internet. Further information is available at the model’s website (CMU-GDI, 2004). For LCA, the EIO-LCA model can be used directly to assess construction, manufacturing and supply chain impacts while resources used during plant operation can be analyzed and added to the analysis (e.g. fuel used over the plant lifetime) (Lave et al., 2000; MacLean et al., 2000a, 2000b; MacLean and Lave, 1998). The model has been successfully applied in many industries (e.g. the analyses of alternative energy technologies, conventional and alternative fuel/propulsion system automobiles, highway and bridge construction options, E-commerce versus conventional commerce) to analyze the life cycle implications of projects, products, and services and to make comparisons among options.

### 1.2.4 Wastewater Treatment Plant LCAs

A small number of conventional LCAs of WWTPs have been completed. For example, van Loosdrecht et al. (1997) attempted to evaluate the overall environmental impacts that could arise due to process changes implemented to decrease the nitrate concentration in the effluent. Their approach consisted of taking mass and energy balances over a plant operating under two different conditions and comparing the results. While conducting a necessary first step which provided clear data, the authors were unable to conclude whether the increased energy requirement of the process change exerted a greater environmental impact than the reduced impact caused by less nitrate in the effluent. Additionally, the capital and operating cost issues were not examined. Furthermore, as indicated by Bagley and Hofmann (1998), a different process change could have been examined that may have provided decreased nitrate at a reduced energy cost.

The majority of the other LCAs conducted on WWTPs have been completed in European countries. Lundin et al. (2000) summarized these studies and compared the environmental loads from large and small scale conventional wastewater systems with separation systems. Their results demonstrate the value of the LCA approach. Most of the important environmental advantages of separation systems only became apparent when the LCA models included a larger set of activities than are traditionally associated with the wastewater system.

The research team for this study completed an assessment of greenhouse gas (GHG) emissions by Canadian wastewater treatment facilities for Environment Canada (Enviromega et al., 2002). Aspects of LCA were used in the study to determine the upstream GHG emissions due to electricity generation used by the facilities during wastewater treatment. The study revealed that significantly different levels of upstream GHG emissions could be expected province-by-province depending on the mix of power generation types (thermal, hydro, nuclear, other) and the types of treatment processes being used within each province (e.g. lagoons, conventional activated sludge, extended aeration).
The studies to date have made an important contribution in focusing stakeholders on the importance of systems level assessments and the usefulness of LCA for informed decision making. However, the true value of LCAs for the wastewater treatment industry as a tool to reduce overall environmental impacts (not just those due to effluent discharge) while also saving money has not been developed.

1.3 Inadequacy of the State of the Art

The economic attractiveness of recovering energy from anaerobic solids treatment depends on a number of factors. Some, such as the anaerobic technology and the energy recovery alternative to be used, have been examined relatively extensively, but independently. Others, such as recovering hydrogen directly from anaerobic treatment and fuel cells for energy have not been thoroughly examined. The capital, operating and maintenance costs of the studied technologies can be determined. However, the economic attractiveness of recovering energy also depends on factors that have not been thoroughly examined. These include the efficiency of the ERS, the cost of alternative energy sources, and the market for excess energy. Additionally, the life-cycle cost of a municipal wastewater treatment facility and its net greenhouse gas contribution (via the energy produced from fossil fuel combustion, say) and other air emissions due to energy recovery have not been factored into the energy cost-effectiveness evaluation. Finally, new technologies for energy recovery have not been examined.

As energy costs due to deregulation and constricted supplies continue to increase, the need to maximize energy recovery cost-effectively will become important. Life cycle costs will also become more important as the public demands ever more sustainable, engineered processes. Currently, although the states of the art in individual fields are now sufficient to develop it, municipal wastewater treatment facilities do not have a tool for effectively evaluating different combinations of anaerobic solids treatment and energy recovery technologies. The work provided by this study will address that inadequacy by integrating the existing information into a useful tool that can be used to determine the most economically attractive method to recover energy from anaerobic sludge treatment.

1.4 Objectives

1.4.1 Project Objectives

Energy recovery from wastewater solids treatment will occur only to the extent that it is economically viable. There are few support and decision-making tools available to wastewater plant management to assist them in determining the feasibility of adopting a digester gas ERS, especially when comparing the economics of the different technologies over their service life. Therefore, the objectives of this project were to:

6. Develop a model for comparing the benefits and costs of alternatives for energy recovery throughout the lifetime of the treatment facility as a function of plant factors and external parameters.

7. Use the model to develop cost-effectiveness response surfaces for existing and novel sludge treatment and energy recovery alternatives.

8. Predict the impact on overall plant operation due to changes in solids treatment and ERS.
9. Predict the most economically viable sludge treatment and energy recovery alternatives for given input sets.

10. Develop a spreadsheet-based tool in U.S.-based and metric unit versions (designated as LCAMER or Life Cycle Assessment Manager for Energy Recovery), that encompasses the model with a user’s manual (the name of the tool is currently known).

1.4.2 Project Scope

The focus of this project was in the development of the technical and economic functions for anaerobic digestion processes, digester gas pre-treatment systems, and digester gas ERS. It was recognized at the outset that this study boundary would not encompass the full cycle of wastewater residuals collection, conditioning and thickening, treatment, dewatering and ultimate disposal. For example, once the solids have been digested anaerobically, some form of ultimate disposal (possibly including dewatering), such as land application, incineration or landfilling, must be implemented. There are significant life cycle costs associated with each of the disposal alternatives. This project did not also include either the conditioning and thickening, or the dewatering and disposal aspects of wastewater solids management, however, to keep the project at a manageable scope. The combination of the energy recovery LCC module provided in this study, with additional modules that estimates the life cycle costs for pre-conditioning and ultimate disposal options, would make a powerful and valuable tool for supporting the optimum decisions of wastewater management for handling wastewater residuals. The development of such a LCC tool for both conditioning/thickening and dewatering/disposal to link with the LCAMER tool developed in this project is highly recommended for future research.
CHAPTER 2.0

METHODOLOGY

2.1 Literature Acquisition and Review

Prior to the development of a keyword list for electronic database searches, the potential options available for recovery of energy from anaerobic digestion of wastewater solids were compiled. A table listing the various digestion technologies, GPS and ERS was created, as shown in Table 2-1. From this master list, the appropriate key words for searching the literature could be extracted.

Table 2-1. Possible Component Options for Energy Recovery from Anaerobically Digested Municipal Wastewater Solids.

<table>
<thead>
<tr>
<th>Anaerobic Digestion</th>
<th>Gas Pretreatment</th>
<th>Energy Recovery</th>
</tr>
</thead>
<tbody>
<tr>
<td>Single-stage mesophilic (35°C or 95°F)</td>
<td>no pretreatment</td>
<td>biogas flaring (no energy recovery)</td>
</tr>
<tr>
<td>Two-stage^1 mesophilic</td>
<td>Moisture removal only</td>
<td>combustion in boiler with thermal energy recovery (steam and/or hot water)</td>
</tr>
<tr>
<td>Single-stage thermophilic (55°C, 131°F)</td>
<td>moisture and H₂S removal</td>
<td>combustion in boiler with thermal energy recovery (hot water) and use of superheated steam in steam turbine/generator for electricity generation</td>
</tr>
<tr>
<td>Two-stage thermophilic</td>
<td>Moisture, H₂S, and particulate removal</td>
<td>combustion in engine generator for electricity generation and thermal energy recovery (exhaust gas and/or hot jacket water)</td>
</tr>
<tr>
<td>Two-phase^2 mesophilic</td>
<td>Moisture, H₂S, CO₂, and particulate removal</td>
<td>combustion in direct drive engine for work recovery</td>
</tr>
<tr>
<td>Two-phase thermophilic</td>
<td></td>
<td>combustion in transportation system engine</td>
</tr>
<tr>
<td>Temperature phased^3 (meso-thermo)</td>
<td></td>
<td>combustion in turbine for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>Temperature phased (thermo-meso)</td>
<td></td>
<td>combustion in microturbine at small facility (&lt;150 kW) for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>Multi-stage^4 mesophilic</td>
<td></td>
<td>phosphoric acid stationary fuel cell for electricity generation and thermal energy recovery</td>
</tr>
<tr>
<td>Multi-stage thermophilic</td>
<td></td>
<td></td>
</tr>
<tr>
<td>multi-stage temperature phased^5</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

“two-stage” denotes the use of two digesters in series; not the separation of fermentation and methanogenesis
“two-phase” denotes the physical separation of fermentation and methanogenesis into separate digesters, such that the growth conditions for each (pH, etc.) is optimized
more than two digesters in series; not all having the same operating temperature range
more than two digesters in series; not all having the same operating temperature range (e.g. meso-thermo-meso)
The PAFC is the only commercially available fuel cell; other fuel cell types that may be used for stationary purposes include the proton exchange fuel cell, the molten carbonate fuel cell, and the solid oxide fuel cell.

The keyword search strategy used to identify citations of interest to the project is outlined in Table 2-2. In each column, the major heading is combined with terms listed below it using the Boolean logic operator “AND”. For example, from the first column, search strings developed were:

- Anaerobic AND sludge AND digestion (in citation titles)
- Anaerobic AND sludge AND treatment (in citation titles)
- Anaerobic AND biosolids AND digestion
- Anaerobic AND biosolids AND treatment

<table>
<thead>
<tr>
<th>Anaerobic AND</th>
<th>anaerobic AND production AND</th>
<th>energy AND recovery AND</th>
<th>economic* AND (keywords)</th>
<th>life-cycle AND cost* AND</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sludge AND digestion (title)</td>
<td>all 16 possible combinations of (methane or biogas) AND (municipal or domestic)</td>
<td>Anaerobic AND sludge</td>
<td>fuel AND cell*</td>
<td>anaerobic AND digestion</td>
</tr>
<tr>
<td>Sludge AND treatment (title)</td>
<td>Anaerobic AND biosolids</td>
<td>gas AND turbine*</td>
<td>anaerobic AND treatment</td>
<td></td>
</tr>
<tr>
<td>Biosolids AND digestion</td>
<td>fuel AND cell*</td>
<td>anaerobic AND digestion</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Biosolids AND treatment</td>
<td>gas AND turbine* (keywords)</td>
<td>anaerobic AND treatment</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Boiler*</td>
<td>Boiler*</td>
<td>Boiler*</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Co-generation</td>
<td></td>
<td>gas AND turbine*</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>fuel AND cell*</td>
<td></td>
<td></td>
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<td></td>
<td></td>
<td>energy AND recovery</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Co-generation</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

After the searches with the various keyword strings, recovered citations were subjected to a filtering procedure to eliminate those that were captured by the search but were deemed irrelevant to this study. The following citation filtering techniques were applied:

- any citations pertaining solely to wastewater treatment were eliminated
- any citations pertaining solely to industrial sludges were eliminated
- any citations pertaining solely to co-digestion of organic fraction of municipal solid waste (OFMSW) and sludge were eliminated
- any citations pertaining solely to aerobic processes were eliminated
- any citations pertaining solely to mobile applications of fuel cells were eliminated
any citations pertaining solely to landfill gas or landfill leachate were eliminated

any pertaining solely to desalination were eliminated

any citations pertaining to combinations of aerobic and anaerobic sludge treatment processes were included

anaerobic sludge digestion processes of any nature (e.g. CSTR, ASBR, fixed-film) were included

2.2 Survey

2.2.1 Questionnaire Development

A detailed data collection template to be mailed to candidate POTW sites was prepared in May 2003 and sent out in early June 2003. The template was designed to cover many aspects of this process, such as the type of treatment processes used; details about digester design (e.g., insulation types); operations (e.g., retention time, volatile solids reductions, gas production and quality); energy use and recovery data; distribution of on-site energy use involving data such as numbers and sizes of pump motors; and capital and operating cost data for the anaerobic digestion and recovery system capital and O&M costs. The survey was detailed, but such data are required to develop the process and costing functions used in the life-cycle assessments, and to develop the software that enables POTWs to determine the optimum energy recovery strategy.

Following analysis of the results of the surveys completed by the treatment facilities, the questionnaire was modified slightly to avoid some redundancy in the questions, and to include data such as whether or not the facility was located in a designated ozone non-attainment area which could influence the choice of energy recovery process. The updated survey form was mailed off in October 2003 to an additional 89 treatment facilities. The survey template is given in Appendix A.

2.2.2 Wastewater Site Visits

At the discretion of the Project Team, site visits were made to 19 facilities in the United States and Canada to follow up with first hand interviews the operation of the digestion and energy recovery technologies identified in the completed survey forms. The site visits fulfilled a number of objectives, including checking the validity of data submitted in the survey forms, filling in missing data, discussing operational aspects of the digestion and energy recovery technologies with those professionals operating the units, and to document the site visits with notes and photographs. Notes from the site visits are included in Appendix B.

2.3 Data Storage and Utilization

Plant data retrieved from the returned survey forms were checked for completeness and if necessary, facility personnel were contacted by telephone or email to update the information. The data were then entered in MS-Access database software. Use of the database permits optimum data extraction by using sort and constraint functions.
Cost information was entered in the currency of the country from which the response originated. For example, if American dollars are entered, the entry will be in the U.S. currency field. The exchange rate between U.S. and Canadian dollars for values in the database reflects the 2003-2004 time frame.

2.4 Estimating Effect of Anaerobic Digestion Process Change on Wastewater Treatment Plant Operation

Few wastewater process changes operate in isolation, with negligible impacts at some other locations in the overall treatment facility. The effect of changing or upgrading one anaerobic process to another (e.g. from mesophilic to thermophilic) was assessed in terms of the effect of digester supernatant on the operation of the liquid treatment train. The impact of the digester supernatant quality and quantity was evaluated in two ways. In the first case, a review of the technical literature was used to develop a spreadsheet model for estimating loadings of solids and nutrients returned to the liquid treatment train. In the second case, wastewater treatment software was used to evaluate the effect of the digester supernatant quality and quantity on liquid process operation. The results of this evaluation are summarised in Appendix C.
CHAPTER 3.0

LITERATURE REVIEW

3.1 Anaerobic Digester Processes

3.1.1 Historical Context of Anaerobic Wastewater Solids Digestion

Anaerobic digestion is a biological process used for the stabilisation of excess sludge (primary and biological sludge) produced during the treatment of municipal wastewaters. This process occurs in the absence of oxygen. From a fundamental stand point, anaerobic digestion process consists of three steps: hydrolysis, fermentation and methanogenesis.

During the hydrolysis step, particulate matters are first solubilized by enzymes. The large soluble organic compounds produced (e.g. lipids, polysaccharides, proteins, etc.) are then converted to their monomers (e.g. fatty acids, monosaccharides, amino acids, etc.). During the fermentation step the monomers are converted to short chain organics acids (e.g. acetic acid, propionic acid, butyric acid, etc.), carbon dioxide and hydrogen. All acids except acetic acid are further converted to acetic acid. The final products of fermentation are therefore acetic acid, carbon dioxide and hydrogen. Fermentation is done by the fermentative bacteria. The latter are also called acidogens or acidogenic bacteria. During the last step of the anaerobic digestion process, acetic acid, carbon dioxide and hydrogen are converted to biogas by methanogens. Biogas is the key end product of sludge digestion and it mainly consists of methane and carbon dioxide.

Historically, the discovering of methane can be attributed to Van Helmont and Volta (FAO, 2006). Van Helmont recorded the emanation of an inflammable gas from decomposition of organic matter in the 17th Century while Volta showed as early as 1776 that the amount of gas produced is a function of the amount of decaying vegetation in the sediments from which the gas is released. The real scientific research on anaerobic digestion probably started between 1804 and 1810 with Dalton, Henry and Davy who established the chemical composition of methane (FAO, 2006). In 1868, Bechamp named the organism responsible for methane production from ethanol and showed that, depending on the substrate, different fermentation products were formed (FAO, 2006).

The concept of anaerobic digestion was introduced around the 1870s with the development of the septic tank. There is a general consensus that attributes the invention of the septic tank to Louis Mourras, a French Engineer who designed the system around 1871 (Builder, 1941). This system is considered as the simplest and oldest anaerobic digester (Jewell, 1987). The concept of using digester gas began as early as 1895, when street lighting in a section of Exeter, England, was fuelled by digesting wastewater (Ward, 1974). During the 1920s Imhoff in
Germany designed a continuous anaerobic digestion system, The Imhoff septic tank. A few years later this system was used in several cities inside and outside Germany.

General discussions of anaerobic digestion operation have appeared at various times in the literature. Buswell (1947) discussed the microbiology and theory of the anaerobic sludge digestion process. In a companion paper, Schlenz (1947) discussed practical guidance on operation of anaerobic digesters (only mesophilic operation was considered at the time), covering topics such as volatile acids and pH; scum and foam control; temperature; feed and withdrawal; and supernatant. A similar review was published by Lohmeyer (1959). Pohland (1962) provided an extensive review of the literature to 1962, offering a good historical perspective of the process to that time. Aspects of the digestion process reviewed included mesophilic and thermophilic operation; control of pH, volatile acids and alkalinity; loading rates and retention times and gas production. Dague (1968) also noted the indicators of the progress of the digestion process, and summarized the factors that might be controlled for satisfactory digester operation, including the balance of bacterial populations, a uniform feeding regime, adequate mixing and contact time, maintaining a uniform temperature, and maintaining a suitable pH. Buhr and Andrews (1977) similarly compiled a detailed literature review of the thermophilic anaerobic digestion process.

Rankin (1948) performed a systematic balance of solids in anaerobic digestion systems, and was among the first to suggest that the combination of volume of raw sludge feed and digester retention time related closely to observed VS reduction, regardless of sludge type (primary, biological or mixed).

Around 1953, full-scale thermophilic digestion for sludge treatment was constructed in Hyperion Treatment plant of the city of Los Angeles (Garber et al. 1975). But, during the energy-abundant 1950’s and 1960’s attention gradually shift from the time-consuming and relatively sensitive anaerobic digestion process towards quicker non-biological, more energy-intensive methods of sludge stabilization (Ward, 1974).

McCarty (1964a,b,c,d) published a series of articles on the fundamentals of anaerobic waste treatment, with individual topics of the chemistry and microbiology of anaerobic treatment, environmental requirements and control; toxic materials and their control in the anaerobic process; and anaerobic process design. Young and McCarty (1969) developed the anaerobic filter concept, while Lawrence and McCarty (1969) developed the mathematics for a kinetic model of methane fermentation during anaerobic treatment. Lawrence and McCarty (1970) developed a simple model to describe the operation of the completely stirred tank reactor (CSTR). The CSTR is currently the most common system utilized for anaerobic sludge digestion either in the conventional North American “pancake” configuration, or as egg-shaped tanks.

Torpey and Melbinger (1967) in New York City experimented with recirculation of digested sludge for overall reduction of volume of treated sludge for disposal. In essence the process involved control of the sludge biomass, a form of SRT control for the anaerobic digester. The optimum reduction of digested sludge volume was observed at a recirculation rate of 40-60% of the feed rate. The concept was also confirmed by Pfeffer et al. (1967) in laboratory-scale work, who demonstrated the advantage of a long retention time due to sludge recycle versus a
short retention time with no recycle when operated at the identical loading rate. Pfeffer (1968) demonstrated that destruction of VS, total gas production and process stability all increased significantly when the digester SRT was increased by digester solids recycle.

The start-up and operation of newly constructed high-rate anaerobic digestion systems at two Chicago plant were reported by Lynam et al. (1967). Start-up could be achieved with Imhoff tank sludge or a lagooned sludge. Stable mesophilic operation was achieved at SRT as low as 10 days.

McCarty (1964a) noted that methane production could be predicted stoichiometrically, by applying a relationship developed by Buswell and Mueller (1952), namely:

\[ C_nH_{a+b} + (n - a/4 - b/2) H_2O \rightarrow (n/2 - a/8 + b/4) CO_2 + (n/2 - a/8 - b/4) CH_4 \]

Andrews (1975a) pointed out that this conversion does not account for the relatively small fraction of organic substrate converted to microbial biomass. Andrews and Graef (1971) developed a dynamic mathematical model for the anaerobic digestion process. Graef and Andrews (1974) and Andrews (1975a,b) described process control strategies for the digestion process, particularly in inhibitory conditions for the methanogens. Control actions suggested included a reduction in the organic loading to the digester, addition of alkaline chemicals, supplementing the digester with well-digested sludge from another source, and scrubbing of carbon dioxide form the digester gas with recirculation to the digester for mixing. Recirculation of digested sludge (SRT control) was also suggested as one of the best strategies for preventing the onset of toxic conditions. Collins and Gilliland (1974) also presented mathematical models depicting the dynamic control of the anaerobic digestion process.

Recognizing the substantial difference in the metabolic characteristics of the acid and the methane formers, some researchers like Pohland and Gosh (1971) has envisioned controlled anaerobic stabilization by phase separation of the two groups in isolated environments. The feasibility of this concept was initially demonstrated with a simple soluble substrate (glucose). Later on, Ghosh et al (1975) demonstrated the feasibility of phase separation on wastewater sludge. Another variant of the two-phase system is the two-stage system experimented by Norrman and Frostell (1977). Massey and Pohland (1978) confirmed that phase separation is feasible and that the process may be applied to both soluble and complex substrates. After conducted an experiment on thermophilic anaerobic digestion of a strong complex substrate, Therkelsen and Carlson (1979) did not recommend thermophilic two-phase scheme except in the case of a rather dilute waste that is rich in starch and protein. In laboratory experiments, Henry et al. (1987) determined that the efficiency of the acidification reactor, based on combined residual volatile acids and biogas, was higher in mesophilic operation than with thermophilic operation. They also found that pH conditions closer to neutral than acidic in an acid-phase reactor resulted in higher concentrations of volatile acids, greater conversion of organic matter, and higher methane yields.

At the beginning of the second century of anaerobic digestion there were many opportunities for innovation in anaerobic treatment (McCarty, 1981). In addition to sludge treatment, the process succeeded to make an important breakout in the area of both industrial and
domestic wastewater treatment. Several high-rate reactors were developed and full-scale plants were built throughout the world. Some of these new systems are the upflow anaerobic sludge blanket (UASB) reactor (Lettinga et al., 1980), the anaerobic baffle (AB) reactor (Bachmann et al. 1982), the anaerobic hybrid (AH) reactor (Guiot and van der Berg, 1984) and the expanded granular sludge bed (EGSB) reactor (De Man et al. 1988). These systems were reviewed in detail by Kalogo and Verstraete (1999).

3.1.2 Anaerobic Digestion Process Configurations

3.1.2.1 Low-rate Anaerobic Digestion

Low-rate anaerobic digestion is the oldest anaerobic stabilization process. It was also originally called standard-rate or conventional anaerobic digestion (WEF, 1995), although the term conventional is out of date. Figure 3-1 shows the schematic of a low-rate digester. Typically, the reaction tank consists of a cylindrical, square, or rectangular tank with a sloping bottom and a flat or domed roof. All of the microbiological reactions of anaerobic digestion happen in this single tank.

Because there is no auxiliary mixing, several layers exist in the tank. Rising gas bubbles during methanogenesis may cause internal mixing. Methane gas that accumulates in the headspace of the tank is collected for storage or use. Scum accumulates on the liquid or supernatant surface. The stabilized solids settle to the bottom for removal and further processing. The supernatant is drawn off and recycled. Between the supernatant and the stabilized solids is the active layer. Grit...
and scum will accumulate on the bottom and top of the tank, respectively, decreasing the effective volume.

Low-rate anaerobic digestion has traditionally been considered only for small WWTPs, under 3450m³/d (0.91-mgd) (WEF, 1995; WEF, 1998). The SRT for low-rate anaerobic digestion is 30-60 days (WEF, 1995), resulting in large tank sizes and high initial construction costs. The digester is intermittently fed with sludge and the volumetric organic loading rates are low, normally, only 0.64-1.60 kgVS m⁻³ day⁻¹ (0.0053-0.013 lb VS/gal-d). There is no auxiliary mixing provided in this system, and, in general, there is no heating system in this alternative. If the digestion rate needs to be increased, an external heat source may be supplied. The application of low-rate anaerobic digestion has been decreasing in recent years (WEF, 1998).

3.1.2.2 High-rate Anaerobic Digestion

High-rate anaerobic digestion systems are characterized by supplemental heating, auxiliary mixing, uniform feeding rates, and sludge thickening before digestion (WEF, 1998). The performance of the anaerobic digester is improved through providing uniform environmental conditions in the digester. Figure 3-2 illustrates a typical high-rate anaerobic digestion system. The volume required for adequate digestion is reduced versus the low-rate system, and the stability of the process is improved.

Heating during digestion increases the growth rate of microorganisms, the digestion rate, and gas production. High-rate systems can be divided into two categories, mesophilic and thermophilic, depending on the temperature maintained in the digester. The sludge in the digester is mixed by gas recirculation, pumping, or draft-tube mixers. For a high-rate digester, uniform feeding is important. The sludge should be pumped to the digester continuously or on a 30-min to 2-hr time cycle to help maintain constant conditions in the reactor.

High-rate mesophilic digestion occurs when the temperature range in the reactor is approximately 30-38°C (86-100°F). The SRT for high-rate mesophilic digestion is typically 12-25 days (Oleszkiewicz and Mavinic, 2001). Compared to low-rate digesters, high-rate mesophilic digesters can be operated at higher organic loading rates, 1.6-3.2 kgVS m⁻³ day⁻¹ (0.013-0.027 lb VS/gal-d) (Metcalf and Eddy, 1991). The gas production rates from mesophilic digestion are typical in the range of 0.75-1.12 m³ biogas per kg VS destroyed (12-18 scf/lb VS destroyed) (Metcalf and Eddy, 2003). The VS reduction is 45-50% (Metcalf and Eddy, 2003), which is higher than the low-rate system, and more gas is produced in the mesophilic high-rate system.

High-rate mesophilic digestion may operate as a single- or two-stage system. Figure 3-2 is a typical single-stage digester. This type of digester involves a single continuous-flow stirred reactor. External heating ensures a proper operating temperature. Auxiliary mixing is also needed. Because the sludge is continuously mixed, there is no supernatant separation in the digester. The digester may have a fixed roof or a floating cover. Floating covers can provide excess gas storage capacity. For a fixed roof, the biogas may be collected and stored in a separate gasholder, at either low-pressure or high-pressure. Digested biosolids are conveyed to be further dewatered or otherwise treated. Two-stage high-rate mesophilic digestion will be discussed in the next section.
High-rate mesophilic digestion is the most common implementation of anaerobic sludge digestion in North America. Most full-scale WWTPs in Canada and the United States that anaerobically digest sludge use mesophilic digestion (Burrowes, 2000). Mesophilic digestion is mostly used for municipalities with sewage flow exceeding 20,000 m³/d (5,283,000 gpd) (Oleszkiewicz and Mavinic, 2001). Compared to thermophilic treatment, there are several disadvantages, including lower VS reduction, lower methane production, and significantly higher fecal coliform contents (Holbrook et al., 2002).

![Diagram of High-rate Anaerobic Digestion](image)

**Figure 3-2. High-rate Anaerobic Digestion.**

High-rate thermophilic digestion occurs when the temperature range in the digester is approximately 50-60°C (122-140°F), which is suitable for the growth of thermophilic bacteria. As with mesophilic high-rate digestion, external heating and auxiliary mixing are both needed for thermophilic digestion. The design criteria and performance of thermophilic digestion are somewhat different from those of mesophilic digestion. The typical SRT for thermophilic digestion is 10-12 days and the volumetric organic loading rate is 3.0-8.7 kgVS m⁻³ day⁻¹ (0.025-0.073 lb VS/gal-d).

Compared to mesophilic digestion, the advantages of thermophilic digestion include higher loading rates and shorter SRT, higher pathogen destruction, higher VS reduction, higher methane gas production, better dewatering characteristics, and smaller digester volumes (Ghosh et al., 1995; Han and Dague, 1997; Schafer et al., 2003). Thermophilic digestion also has disadvantages compared to mesophilic digestion, such as higher heating cost, higher volatile fatty acids (VFAs) concentrations in the effluent, pronounced odors from increased production of C⁵ and C⁶ volatile acids, higher ammonia levels, difficulty in drawing off supernatant, and susceptibility to upsets from shock loading (Ghosh et al., 1995; Han and Dague, 1997; Holbrook et al., 2002).
High-rate thermophilic digestion can also be classified into single- or two-stage thermophilic digestion, and staged thermophilic digestion as well. Single-stage thermophilic digestion is the simplest version of the thermophilic high-rate process. Figure 3-2 is also a typical schematic of this alternative. Several large cities in the United States and Canada, including Los Angeles, California (Witzgall et al., 2003), New York (WEF, 1998), and Vancouver, BC (Witzgall et al., 2003), have used thermophilic digestion. Several plants in North America now use staged thermophilic digestion to improve performance (Schafer et al., 2003). Two-stage thermophilic digestion and staged-thermophilic digestion will be discussed in the next section.

3.1.2.3 Two-stage Anaerobic Digestion

Historically, a two-stage digestion system has involved a primary digestion reactor that is heated and completely mixed, and a secondary tank that is neither heated nor mixed. The two stages are arranged in series (Figure 3-3). The first stage serves as a high-rate single-stage system, in which the temperature is controlled at either mesophilic or thermophilic range. Most of the gas is produced from the first stage reactor. The second stage provides an additional “safety” capacity. Its major function is solid-liquid separation, in which no mixing is provided, thus allowing for biosolids concentration (gravity thickening). The solid-liquid separation produces a supernatant layer in the reactor. The second stage also may serve other functions, such as providing storage capacity and reduced short-circuiting of the overall process.

![Two-stage Anaerobic Digestion](image)

Figure 3-3. Two-stage Anaerobic Digestion (unmixed in 2nd stage).

This type of two-stage digestion is effectively equivalent to a single-stage high-rate digestion. For this two-stage anaerobic digestion system, the first stage of the system should provide enough SRT to ensure normal, reliable methane gas production under all conditions (WEF, 1998; Schafer and Farrell, 2000). However, anaerobically digested solids may not settle well in the
secondary tank, resulting in a high concentration of suspended solids in the supernatant. This traditional two-stage digestion is seldom used in modern design (Metcalf and Eddy, 2003).

In recent designs, two-stage digestion operates the secondary reactor as a mixed reactor. One function of the secondary reactor is to serve as a holding tank for digested biosolids before dewatering or other follow-on processes. Figure 3-4 provides the schematic of this two-stage digestion. Both stages should be mixed continuously, and gas is collected from both stages (Oles et al., 1997; Schafer and Farrell, 2000). This type of two-stage digestion differs from single-stage digestion because methanogenesis occurs in both stages.

Figure 3-4. Two-stage Anaerobic Digestion (mixed in 2nd stage).

Configurations of two-stage digestion include two-stage mesophilic digestion and two-stage thermophilic digestion. Auxiliary mixing and heating are provided to the secondary reactor, as well as the heating to maintain the proper temperature. Two-stage mesophilic digestion is the simplest type of staged anaerobic digestion. From 1999 to 2000, the Hyperion WWTP in Los Angeles (California, U.S.) operated two-stage mesophilic digestion, which consisted of egg-shaped digesters and conventional digesters. The SRTs for the first stage and second stage were 10 days and 8 days, respectively. The VS reduction reached 60-64 % (Schafer and Farrell, 2000; Witzgall et al., 2003). Other full-scale data for two-stage mesophilic digestion are scarce. The limited data indicate that two-staged mesophilic digestion reduces odors (primary benefit); improves VS Reduction by a few percentage points, depending on the system’s configuration; increases the gas production slightly; and reduces short circuiting of solids in the system (Schafer and Farrell, 2000).

Two-stage thermophilic digestion is applied at Lions Gate WWTP in Vancouver (British Columbia, Canada). This configuration can meet Class A biosolids requirements unlike the single-stage thermophilic digestion previously used at the plant (Shimp et al., 2000b). The Hyperion Plant, mentioned before, changed to two-stage thermophilic digestion from its previous mesophilic mode in late 2000. The SRT of the system was changed to 13 days and the VS Reduction reached 59.7%. The process could produce Class A biosolids (Witzgall et al., 2003).
3.1.2.4 Temperature-phased anaerobic digestion (TPAD)

Temperature-phased anaerobic digestion (TPAD) is also a two-stage digestion process. TPAD combines in one system both mesophilic and thermophilic digestion process stages connected in series. The aim is to improve the stabilization process as much as possible. Both of the stages are heated and mixed for proper environmental and process conditions. In each temperature phase, there is no requirement for metabolic phase separation. Hydrolysis, acidogenesis and methanogenesis occur and balance in each reactor.

There are two different typical configurations of temperature-phased anaerobic digestion. The most common implementation of TPAD uses thermophilic digestion (with a temperature range of approximately 55°C or 131°F) as the first phase, followed by mesophilic digestion (in the 35°C or 95°F temperature range) (Figure 3-5). This configuration is a process patented by Iowa State University (Schafer and Farrell, 2000) and is designed to produce a final product with minimal volatile acid and odor levels, and reduce the effects on polymer conditioning for solids dewatering in the system (Schafer et al., 2003).

Vandenburgh and Ellis (2002) showed that the thermophilic reactor in the TPAD process performed more than two-thirds of the VS destruction and nearly all the fecal coliform destruction. The mesophilic digester’s purpose was to provide final polishing to reduce the intermediate by-products from the thermophilic reactor.

Temperature-phased anaerobic digestion can combine the advantages of thermophilic and mesophilic digestion, and appears to avoid the disadvantages of each. The advantages of TPAD are (Han and Dague, 1997; Holbrook et al., 2002):

- Significantly improved VS Reduction;
- Increased methane production;

![Figure 3-5. Temperature-phased Anaerobic (Thermo+Meso) (TPAD).](image-url)
Minimized odor production associated with thermophilic digestion;

Increased pathogenic bacteria reduction and potential to produce Class A biosolids;

More stable biosolids and improved dewatering characteristic of biosolids;

Less volume required for same degree of VS Reduction; and,

Greater capacity to absorb shock loading than conventional anaerobic digestion.

The disadvantages are (Han and Dague, 1997; Vandenburgh and Ellis, 2002)

Higher energy required for thermophilic phase;

Higher ammonia content in the effluent; and,

Greater odor problems during thermophilic phase.

Oles et al. (1997) in his study suggested that the design SRT for TPAD (thermophilic + mesophilic) is 2-3 days and 12-15 days for thermophilic and mesophilic, respectively, after the investigation of ten full–scale TPAD plants. Schafer and Farrell (2000) suggested that the SRT for thermophilic and mesophilic stages are 3-8 days and 10-12 days, respectively. Longer SRTs are required to ensure stable operation in full-scale plants (Schafer et al., 2003).

TPAD has presently been applied to ten systems in Germany (Oles et al., 1997). There are also eight TPAD plants now operating in the US, most of which are in the Midwest (Schafer and Farrell, 2000). For example, the 20,800 m$^3$/d (5,495,000 gpd) Papillion Creek WWTP (Omaha, Nebraska) is one of the largest temperature-phased digestion systems in the country. Significant improvement of VS Reduction (15-20%) has been achieved through converting mesophilic single-stage digestion to TPAD (Schafer and Farrell, 2000).

Another configuration of temperature-phased digestion system is organized in a mesophilic-thermophilic configuration, in which mesophilic digestion precedes the thermophilic digestion (Figure 3-6). For example, the Metropolitan Water Resources District WWTP in Chicago (Illinois, U.S.) used the mesophilic-thermophilic configuration of temperature-phased anaerobic digestion and their VS Reduction achieved 55% overall (Schafer and Farrell, 2000).
There are also other modes of TPAD. A three-stage TPAD (Figure 3-7) is used in the 227,000 m³/day (60 mgd) Village Creek WWTP in Birmingham (Alabama, U.S.) as reported by Holbrook et al., (2002). Operation consisted of three phases, a thermophilic stage operated at 57°C to 60 °C with a design SRT of 8.2 days and an organic loading rate of 4.0 kgVSS m⁻³ day⁻¹ (0.033 lb VSS/gal-d), followed by mixed and heated fixed-cover mesophilic digesters with a design SRT of 12.3 days, and finally floating-cover unmixed and unheated mesophilic digesters with a design SRT of 12.3 days. The total design SRT is 32.8 days (Holbrook et al., 2002). The performance of this process reached, as expected, more than 58% of VS Reduction, increasing methane production enough to heat the digesters, and producing Class A biosolids. This digestion process is reported to be easily operated, stable, and generally odor-free (Holbrook et al., 2002).

Figure 3-7. Design of Temperature-phased Anaerobic Digestion in Village Creek WWTP.
3.1.2.5 Two-phase Anaerobic Digestion (Acid-gas phased Anaerobic Digestion - AGAD)

Two-phase anaerobic digestion is also a two-stage process, where the first stage is an acid phase reactor and the second stage is a methanogenic reactor (Figure 3-8). A two-phase system involves a physical separation of acidogenesis and methanogenesis, and enriches different bacteria in each digester by independently controlling the digester conditions, thereby improving reaction kinetics and stability. The first phase (acidogenesis) is operated to optimize hydrolysis acidogenic growth, and the second phase (methanogenesis) operated to optimize methanogenic growth. The first stage has a pH of 6 or less and a short SRT (1-2 days), which maximizes acid production and minimizes methane and total gas production. The second stage has a neutral pH and a longer SRT (10-14 days) to allow growth of methanogenic organisms (Murthy, 2001). The organic loading rate is the major concern of the system. The organic loading rate can be 32 kgVS m$^{-3}$ day$^{-1}$ (0.27 lb VSS/gal-d) in the acid-phase while the overall organic loading rate is much lower, 3.2 kgVS m$^{-3}$ day$^{-1}$ (0.027 lb VSS/gal-d) (Murthy, 2001).

![Figure 3-8. Two-phase Anaerobic Digestion (Acid-gas Phased Anaerobic Digestion).](image)

The potential advantages of a two-phase anaerobic digestion system are:

- Optimized growth of hydrolysis-acidogenic and coupled acetogenic-methanogenic bacteria in the two-phase process through physical separation (Ince, 1998);
- Improved overall process stability and control;
- Shorter SRTs and reduced overall volume (Schafer and Farrell, 2000);
- Higher organic loading rates;
♦ Increased specific activity of methanogens leading to a higher methane production rate;

♦ Increased overall COD and volatile solids reduction efficiencies, which can reach as high as 50-60 %;

♦ Higher pathogen reduction rate and potential to produce Class A biosolids (Schafer et al., 2003).

The disadvantages of phase separation are reputed to include: 1) hydrogen build-up in the first-phase reactor during the acid formation to levels inhibitory to acid-producing bacteria (Bhattacharya et al., 1996), 2) loss of potential methane formation from H₂ and CO₂ in the acidogenic step, and 3) elimination of possible interdependent nutritional requirements of acid and methane formers (Conrad et al., 1985; Fox and Pohland, 1994).

The two-phase system involves either phase (acid or gas phase) at mesophilic or thermophilic temperature. The configurations can be mesophilic-mesophilic, mesophilic-thermophilic, and thermophilic-mesophilic. Both phases at mesophilic temperatures are the most common implementation (Schafer et al., 2003), but not sufficiently reliable to produce Class A biosolids (Gray et al., 2003). The performance of mesophilic/thermophilic two-stage digesters was reported to be better than that of thermophilic/mesophilic two-stage digesters, and the digested biosolids from these two configurations could meet Class A fecal coliform requirements (Gray et al., 2003).

Two-phase acid-gas anaerobic digestion (AGAD) has been implemented only in the United States. A full-scale two-phase anaerobic digestion system is the Woodridge-Green Valley WWTP, which is operated by municipalities in DuPage County (Illinois, U.S.). The configurations in this plant are mesophilic-thermophilic and mesophilic-mesophilic.

3.1.2.6 Three- and Multi-staged Anaerobic Digestion

WWTP managers in North America are considering using advanced anaerobic digester technologies to improve the efficiency of anaerobic digestion, reduce the volume requirements, and improve pathogen reduction to meet Class A requirements to be land applied without restrictions. Three-/Multi-staged anaerobic digestion, one type of advanced anaerobic digestion technology, is a new emerging alternative, in which at least three stages exist. Process design and operation become more complex as the number of stages increases.

One three-phased digestion process combines acid-gas phased digestion and TPAD. The schematic of this process is shown in Figure 3-9. The Inland Empire Utilities Agency has implemented this three-phased digestion (Schafer et al., 2003). The first phase is an acid tank, in which hydrolysis and acidification happen under short SRTs (3 days). Following the first stage is a TPAD system, which is a configuration of thermophilic and mesophilic temperatures. The SRTs of both reactors are 12-14 days and 14-16 days, respectively. In the actual operation, the third phase of the system at Inland Empire is not operated at a specific temperature, in which the product is allowed to reach its own temperature from radiant heat loss to the environment. Data from this plant shows that the process provides some improvement in VS Reduction, from 54% in
mesophilic digestion to 56%. The primary benefits are the improvement of mechanical dewatering characteristics of the product and the elimination of odor problems (Schafer et al., 2003).

Figure 3-9. Combination of 2-phase and TPAD Anaerobic Digestion System.

Two or more thermophilic reactors are set up in series to improve VS Reduction and meet the Class A criteria. This configuration is termed staged or extended thermophilic anaerobic digestion (Murthy, 2001). Two-stage thermophilic digestion was discussed in the previous section. Three-or more staged thermophilic digestion is rare in North America. Because only a few facilities have implemented this process, the operation data about this alternative is limited.

The OWASA Mason Farm plant in North Carolina has significantly improved VS Reduction from 43% to 65% by switching from mesophilic digestion to a three-staged thermophilic digestion. Before changing, the plant had difficulty consistently meeting Class A biosolids requirements. The SRTs for the three stages are 19, 10, and 10 days, respectively (Schafer et al., 2003).

One example of four-staged thermophilic digestion is the 500,000m$^3$/day (132 MGD) Annacis Island plant in Vancouver (British Columbia, Canada) which implemented this design in the mid-1990s. The SRTs for the second, third, and fourth stages are approximately 2.5 days each and the SRT for the first stage is 17-18 days. The total system SRT is approximately 25 days (Ahring et al., 2002). The VS reduction of the system is approximately 63%, a significant improvement over non-staging. The fecal coliform data showed that the final biosolids product consistently achieved low densities Class A biosolids requirements (Holbrook et al., 2002; Schafer and Farrell, 2000; Witzgall et al., 2003).
3.1.2.7 Comparison of Anaerobic Digestion Technologies

Alternatives Investigated in This Project.
The anaerobic digestion alternatives investigated in this project are as follows.

♦ High-rate mesophilic anaerobic digestion;
♦ High-rate thermophilic anaerobic digestion;
♦ Two-stage anaerobic digestion (including mesophilic and thermophilic);
♦ Temperature-phased anaerobic digestion;
♦ Two-phase (acid-gas phased) anaerobic digestion;
♦ Three-/Multi- phased anaerobic digestion.

Mesophilic anaerobic digestion has been widely used to stabilize primary sludge since 1950 around the world (Ahring et al., 2002; Nielsen and Petersen, 2000). In order to find low cost methods to achieve Class A biosolids and continue land application, thermophilic anaerobic digestion has become an attractive alternative, and many plants have changed to thermophilic digestion from mesophilic digestion (Gray et al., 2003; Nielsen and Petersen, 2000; Zabranska et al., 2000).

The alternatives of temperature-phased anaerobic digestion, acid-gas phased anaerobic digestion, thermophilic anaerobic digestion, and three-/multi-phase anaerobic digestion can be termed as advanced anaerobic digestion processes. Many POTWs are now considering advanced anaerobic digestion processes in order to: 1) increase digester loading so they can decrease the size and cost of new facilities; 2) avoid expanding existing facilities; 3) reduce the investment and operation cost; and 4) improve the performance of anaerobic digestion by increasing VS Reduction and gas production and reducing the amount of biosolids (Shimp et al., 2000a).

Advantages and Disadvantages of Each Alternative: Each anaerobic digestion alternative has advantages and disadvantages. Mesophilic high-rate anaerobic digestion is the standard (conventional) anaerobic digestion. The advantages and disadvantages of others investigated anaerobic digestion alternatives compared to mesophilic high-rate digestion can be found in Table 3-1.

3.1.2.8 Digester Tank Configuration
There are two principal tank configurations (shapes): conventional- and egg-shaped digesters. The conventional-shaped digesters are the traditional shape of the digesters, which is the most common shape used in the U.S. (Metcalf and Eddy, 2003). The advantages of conventional-shaped digester include: can accommodate shape for gas storage; able to be equipped with gasholder covers; a low profile (large radius to height ratio); able to use conventional construction techniques; and less expensive to construct compared to egg-shaped digesters.
The disadvantages of the conventional-shaped digesters are inefficient mixing and dead spaces, which results in grit and silt accumulation; large liquid-gas surface area resulting in scum and foam formation; and higher operation costs.

Table 3-1. Advantages and Disadvantages of Investigated Anaerobic Digestion Technologies as Compared to Mesophilic High-Rate Anaerobic Digestion.

<table>
<thead>
<tr>
<th>Technology</th>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
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</table>
| Mesophilic high-rate anaerobic digestion | Conventional process  
Non-proprietary  
Proven track record in WWTPs  
Most widely implemented process across North America | Poor dewatering characteristics as compared to raw solids dewatering  
Low VS Reduction  
Potential foaming problems  
Longer SRT to achieve desirable VS Reduction compared to the following technologies |
| Thermophilic high-rate anaerobic digestion | Increased reaction rates, smaller digester volumes  
Improved VS Reduction  
Higher gas production  
Decreased foaming problems  
Increased pathogen destruction  
May produce Class A biosolids | Higher operation costs  
More offensive odors  
More energy for heating |
| Temperature - phased anaerobic digestion | Relatively simple to convert from existing multiple tank system  
Robust anaerobic digestion process  
Improved VS Reduction  
Requires less reactor volume for same level of VS Reduction  
Improved gas production  
May produce Class A biosolids  
Control of odors | Patented process (Iowa State University)  
Higher ammonia levels  
Produces odorous biosolids during thermophilic digestion  
Limited use in North America  
Limited operation data available  
May require more energy |
| Two-phase anaerobic digestion | Reduced foaming problem  
Increased gas production  
May improve dewaterability of biosolids  
Improved VS Reduction  
May produce Class A biosolids  
Greater system stability | Requires more energy for thermophilic temperature if thermophilic stage is applied  
Produces higher ammonia levels  
Limited use in North America  
Limited operation data available  
Produces odorous biosolids during thermophilic digestion if thermophilic stage is applied |

Because anaerobic bacteria which survive under mesophilic conditions (35°C, 95°F) are sensitive to variations in the environment, uniform conditions must be maintained to promote healthy biological growth and activity in the digester. Conventional digesters may have difficulty maintaining such conditions because of their relatively flat profile and large surface area (Brinkman and Voss, 1999).

Egg-shaped digesters were developed in Germany to address many of the problems inherent in conventional digesters (Brinkman and Voss, 1999; Metcalf and Eddy, 2003). Egg-shaped digesters have several advantages, including: minimum grit and debris accumulation, reduced scum and foam formation, high mixing efficiency, low operating costs, and efficient land use (less land per unit volume) (Brinkman and Voss, 1999; Metcalf and Eddy, 2003; Witzgall et
The results of a survey in Germany about egg-shaped digesters showed that mixing energies for egg-shaped digesters are typically only 40-60% of those applied to conventional digesters (Brinkman and Voss, 1999). A number of egg-shaped anaerobic digesters are now operating in North America.

The disadvantages of egg-shaped digesters include: higher construction cost per unit volume than conventional-shaped digesters (Witzgall et al., 1998); and the potential for higher heating loss due to the material of digesters. The material used to construct egg-shaped digesters is steel. Insulation and supplemental heating should be used because heat loss through a steel vessel can be ten times (or more) greater than through a concrete one.

3.1.3 Effect of Supernatant Quality from Digester Process Conversion on Liquid Process Train

The effect of anaerobic digester supernatant on the operation of liquid treatment processes has been recognized for some time. In a paper published by the Water Environment Federation, the U.S. EPA (1987) presented a Design Information Report on sidestreams in WWTPs. The principal contaminants documented for anaerobic digester supernatant were the biochemical oxygen demand (BOD₅) and total suspended solids (TSS). The range of concentrations reported for these contaminants was 100 - 2000 mg/L (0.0008 – 0.017 lb/gal) for BOD₅ and 100 – 10,000 mg/L (0.0008 – 0.084 lb/gal) for TSS. Total BOD₅ concentration in digester supernatant is the sum of soluble BOD₅ and BOD₅ of the suspended solids. What is less clear is the effect of changing from one anaerobic technology, such as high-rate mesophilic digestion to one of the more innovative anaerobic technologies. Changing supernatant quality will have the greatest effect on facilities operating at or near design capacity. Those with excess capacity can usually absorb the sidestream loadings, whereas those with no excess capacity may experience a decline or failure in process performance U.S. EPA (1987). Strategies to handle digester supernatant include operating at higher mixed liquor concentrations and aeration rates. If the plant secondary clarifier settling capacity, return sludge pumping and aeration systems are not designed to accommodate this strategy, however, it may be best to treat the supernatant in a separate tank.

Daigger (1998) notes that the most significant effects of anaerobic digester supernatant and other process sidestreams on liquid process operation are due to increased loadings of ammonia-nitrogen, often as periodic discharge. Supernatant return may be most deleterious to operation during high flow periods (e.g. weekday operation in the early afternoon). For treatment facilities that nitrify or partially-nitrify, the excess ammonia load may result in ammonia-N breakthrough, or depletion of alkalinity in the influent wastewater, causing pH decline in the aeration tank.

Anaerobic digestion of biological nutrient removal (BNR) sludges can also contribute high loadings of ammonia-N and phosphorus to the liquid process treatment (Jeyanayagam and Husband, 2002). Anaerobic digestion of BNR sludges can release up to 130 mg/L (0.0011 lb/gal) of total P and 1,000 mg/L (0.0083 lb/gal) of ammonia-N (Pitman et al., 1991). Periodic return of sidestreams such as anaerobic digester supernatant can nutrient spikes to the liquid treatment system, causing:

♦ Inadequate aeration basin volume and/or aerator/blower capacity to achieve reliable nitrification over 24 hours;
Lowered denitrification rate because of a decreased COD/TKN ratio (lack of adequate substrate for nutrient spike condition);

Reduced excess phosphorus uptake because of a decreased COD/TP ratio (lack of readily biodegradable substrate for nutrient spike condition); and

Nitrification inhibition due to lower mean cell retention time caused by solids recycled to the aerobic reactor.

Appleton et al. (2003) used the BioWin32 simulator developed and commercialised by EnviroSim Associates in Dundas (Ontario, Canada) to assess the affect of decommissioning a thermal sludge conditioning unit on overall plant performance. Reductions in aeration due to decreased BOD₅ and ammonia-N loadings were predicted, but effluent limits were jeopardized because the two-stage nitrification system was affected by inadequate organic loading to the second stage. Bypass of part of the flow to the second stage was recommended, with conversion of part of the existing secondary stage to an unaerated selector.

As indicated in the methodology section, wastewater treatment software (BioWin32) was used during this study to evaluate the effect of the digester supernatant quality and quantity on liquid process operation. The results of this evaluation are summarized in Appendix C.

3.2 Digester Gas Pretreatment Systems

When digester gas is used as fuel, several trace compounds have the potential to damage the ERS, causing service interruptions and requiring repairs (Schweigkofler and Niessner, 2001). Among the digester gas constituents, water (H₂O), CO₂, hydrogen sulfide (H₂S) (Constant et al., 1989) and siloxanes (Tower, 2003b) are known to have corrosive effects on ERS.

3.2.1 Moisture

Gas exiting the digester is saturated with water. Any decrease in gas temperature will result in condensation and subsequent corrosion of ERS. One way to alleviate this is to maintain gas above dew point between the anaerobic digester and the ERS (Chambers and Potter, 2002). This is not a possibility for WWTPs that use ERSs preceded by compression or cooling stages. Other WWTPs elect to remove water vapor to obtain the additional benefit of removing other compounds as well (discussed later).

The four main systems of removing water vapor are through use of an accumulator or drip tray; a desiccant dryer; a refrigeration system; or a condenser. Accumulators are typically located directly downstream of the digester to settle out moisture and solids by providing an enlarged piping area that reduces the gas flow velocity (WEF, 1990). Drip tray assemblies are often located throughout the system at low points in the piping (WEF, 1995). Accumulators and drip trays have low water vapor removal efficiencies so cannot be the only form of moisture removal for ERSs sensitive to water vapor such as microturbines.

Desiccant dryers have higher water vapor removal efficiencies than accumulators. Desiccant dryers generally consist of a vessel or tower loaded with a desiccant material that
absorbs water. The digester gas enters through the lower half of the vessel, and moves upwards, which reduces its velocity as it moves through the unit. These changes cause water vapour to condense out of the digester gas into a “claim area”. The gas continues to move upwards through a bed of drying tablets, which absorb and form a solution with the water. Desiccant dryers can be operated without any energy input, with the exception of regenerative desiccant dryers. When the desiccant regeneration is required, the process may require a small volume of compressed air or heated atmospheric air from a blower to dry the desiccant (Air & Vacuum Process, Inc., 2004).

The Lewiston Water Pollution Control Plant (New York, U.S.) uses two tower dryers manufactured by Van Air to dry digester gas. A large one is placed upstream to an activated carbon unit and a smaller one is placed immediately upstream to gas compression for use in a microturbine.

Refrigeration chiller systems and condensers can also be used but these can become energy intensive, particularly if the volumes of gas treated are large. Joint Water Pollution Control Plant (JWPCP) in Carson (California, U.S.), and Annacis Island WWTP in Vancouver (British Columbia, Canada) are examples of plants that use refrigeration as a part of a larger gas pretreatment train (see siloxane treatment section for more information).

3.2.2 Siloxanes

Siloxanes are organic silicon polymers manufactured as additives that improve the properties of consumer products such as cosmetics, hair care products and deodorants. They currently are not regulated in the U.S. because they do not contribute to air pollution (CAT, 1997a).

Minute quantities of siloxanes enter the wastewater system, and are sorbed onto solids pumped to the digesters. As the sludge is heated in the digesters, the siloxanes are desorbed from the solids and volatilize with the digester gas produced. The siloxanes found in digester gas fall into two categories: linear and cyclical (Tower, 2003b). Common linear siloxanes are hexamethyldisiloxane (L2) and octamethyltrisiloxane (L3) while octamethylcyclotetrasiloxane (D4) and decamethylcyclopentasiloxane (D5) are the most common cyclical siloxanes found (Tower, 2003b; Huppmann et al., 1996; Schweigkofler and Niessner, 1999).

There appears to be a correlation between thermophilic digestion of wastewater sludge and higher siloxane levels in the digester gas. Observations of the Annacis Island WWTP have found siloxane-induced problems with the ERSs after a thermophilic digester was implemented (Annacis Island WWTP, 2004). Gary et al. (2001) also observed a 60-300% increase in siloxane concentrations in thermophilic digesters from the concentration found in mesophilic digesters.

3.2.2.1 Impact on Energy Recovery Systems

Once in the digester gas, the siloxanes flow into the ERS where, with the exception of fuel cells, the digester gas is combusted to produce useful energy. The organic part of siloxanes is oxidized, leaving silicates and micro-crystalline quartz behind. Silicates and micro-crystalline quartz strongly bond to the heated metal surfaces of digester gas ERS leading to the need for
frequent and expensive maintenance (Huppmann et al., 1996; Schweigkofler and Niessner, 2001; Tower, 2003b).

When siloxanes are present in the digester gas used by WWTP boilers, deposits often form in boiler tubes reducing its heat transfer efficiency. To remove these deposits, boilers often need to be cleaned or replaced (Tower, 2003a). In internal combustion engines, siloxane-related problems include fouling in the combustion chamber, as deposits often form on the valves, valve seats, piston crowns, and cylinder walls (Tower, 2003b; Schweigkofler and Niessner, 2001). Sometimes these deposits collect under the exhaust valves resulting in burnt valves; this phenomenon reduces compression and engine efficiency (Tower, 2003b).

When gas turbines are used for energy recovery, deposits from the combustion of siloxanes form in the hottest areas, which are mainly the first few rows of nozzles and blades. Prolonged operation of gas turbines with digester gas containing siloxanes can lead to severe erosion of the turbine blades and a sharp drop in operating efficiency (Tower, 2003b).

For the combined-cycle turbine system at the JWPCP, silicate and micro-crystalline quartz deposits caused fouling to occur in the tubing on the heat recovery steam generator unit and on the gas turbine blades (Gary et al., 2001). The tube fouling in the heat recovery system unit was of particular concern because the heat transfer efficiency was decreased. This resulted in an additional 340kg/hr (750 lb/h) of steam generation capacity loss (Gary et al., 2001).

Fouling caused by silicates and micro-crystalline quartz deposits also occurs in add-on air pollution control equipment, and has been observed in at least 15 WWTPs in the U.S. (Glus et al., 1999). Selective catalytic reactor (SCR) catalysts in particular are highly susceptible to siloxane-related damage (Tower, 2003a).

3.2.2.2 Siloxane Tolerances

Most surveyed WWTPs do not remove siloxanes and have yet to report problems. Some WWTPs, however, have observed fouling in their ERSs and air pollution control equipment when operating with digester gas with much lower siloxane concentrations than the limits previously stated. Applied Filter Technology (AFT), the manufacturer of the patented activated carbon-based siloxane removal system, Selective Active Gradient™ (SAG™), claims moderate damage can occur with siloxane concentrations as low as 50 ppb (AFT, 2004). The SAG™ system is a porous pelletized or granular carbon developed in 1996 to remove siloxanes and other contaminants (Tower, 2003b).

Liang et al. (2002) reported that newer internal combustion engines tend to be less tolerant to digester gas contaminants, therefore WWTPs with older units without air pollution control systems do not report fouling (Liang et al., 2002).

Schweigkofler and Niessner (2001) have stated that several engine manufacturers have imposed a limit of 15mg/m³ (0.00094 lb/scf). Assuming the siloxane is compromised of either 100% D4 (296 g/mol) or 100% D5 (370 g/mol) (HPAL, 2005), this is in the range of 0.9 to 1.1 ppm. This is in line with the maximum tolerance limit given by Caterpillar Incorporated for their engines. They give a maximum limit of 0.56mg/MJfuel (1.30 x 10⁻⁶ lb/Btu_{fuel}). For digester gas that
is 60% methane, and assuming the siloxane content is D4 or D5, the siloxane limit is in the range of 0.7 to 0.9 ppm. These limits agree with observations at the Bergen County WWTP, where fouling in the internal combustion engine catalytic oxidizers was found with siloxane concentrations of 2 to 4ppm (only D4 and D5 were measured) (Liang et al., 2002).

### 3.2.2.3 Siloxane Removal Techniques

There are several approaches for removing siloxanes from digester gas. According to Glus et al. (1999), refrigeration, liquid adsorption, and activated carbon are the only technologies that have been used in full-scale biogas application. Of the three techniques, carbon adsorption is the only technology operating full-scale to treat digester gas.

**Condensation:** Chilling to remove moisture is one way to effectively remove some siloxanes from the digester gas (Schweigkofler and Niessner, 2001). Chilling is viable in cases where siloxane concentrations are very high (Tower, 2003a), however, chilling below 3°C (38°F) produces diminishing returns for siloxanes removal and can lose some methane in the condensate (Tower, 2004).

The siloxane removal efficiency of a three-stage condenser system is approximately 80-90% (Glus et al., 1999). However, others estimate that condensation is only able to remove between 10% and 60% of the siloxanes, depending on the species present and other gas constituents (Tower, 2004). According to the supplier of the SAG™ system, chillers should only be used when it is economically justifiable to reduce siloxanes prior to pretreatment of digester gas in the SAG™ system (Tower, 2003a).

The JWPCP is one plant that has decided to carry out initial siloxane reduction using a condensation system. The condensation system selected was a High Pressure/Low Temperature (HPLT) system. The JWPCP has a digester gas flow rate of 245,000m³/day (6000scfm), a pressure of 25.8atm (365psig), and a temperature of 43°C (110°F). The HPLT system cools the digester gas, which contains 4.7ppm of siloxanes, to 4°C (40°F). The result is just over 50% of siloxane removal, leaving the digester gas with a siloxane concentration of approximately 2ppm (Gary et al., 2001).

A pilot scale test was performed on the JWPCP digester gas downstream of the HPLT system to further reduce siloxane concentration. The test involved chilling the gas to -28°C (-19°F) and resulted in an additional 80% of the D4 and D5 siloxanes removal. When a 1-micron filter was added, nearly 100% removal was achieved (Gary et al., 2001).

**Liquid Absorption:** Laboratory studies suggest that siloxane removal through liquid phase absorption, using 48% sulfuric acid at a temperature of 60°C, could remove more than 95% of siloxanes in dry digester gas (Schweigkofler and Niessner, 2001). Similar removal rates were obtained using concentrated (65%) nitric acid. Gas drying is a prerequisite for liquid absorption. Liquid absorbents have been used by some landfills to pretreat landfill gas prior to use in ERSSs in Europe and US SELEXOLO manufactured by Union Carbide is a popular type, and is capable of removing H₂S and CO₂, as well (Glus et al., 1999).
**Activated Carbon Filter:** The studies performed by Liang and Sheehan (2002) and Enochs et al. (2003) evaluated emerging technologies for siloxanes removal such as activated carbon filter systems, synthetic resins, condensation/refrigeration, liquid absorbents, and membrane technology. Both studies concluded that activated carbon filters are best siloxane removal system available because of their ease of implementation, performance capabilities, operational history, installation costs, and maintenance requirements.

Activated carbon has exceptional adsorption properties and large surface area that make it an ideal filter material for a wide range of organic substances in both gas and liquid media (Glus et al., 1999). Activated carbon filters are generally comprised of a vessel containing a bed of activated carbon material through which the digester gas flows. Siloxane and other volatile organic compounds (VOCs) molecules adsorb to the activated carbon pores.

One benefit of activated carbon systems is that the spent material is non-hazardous and can be disposed of in landfills or regenerated. The activated carbon can be regenerated to a high capacity using microwave technology (Liang et al., 1999). Annacis Island WWTP in Delta (British Columbia, Canada) opted for off-site regeneration because their monthly carbon consumption would not justify the additional complexity, maintenance costs and safety concerns of adding a regeneration system on-site (Slezak et al., 2002).

There are many types of activated carbon filters that are capable of removing siloxanes. Pilot tests that were conducted for the JWPCP found solid graphite-based and coconut shell-based type filters have acceptable adsorption capacities before and after regeneration (Gary et al., 2001). However, both filter types had problems with short bed life caused by adsorption of other VOCs in the activated carbon (AFT, 2004).

AFT claims the SAG™ technology can reduce siloxanes concentrations to less than 20ppb, and that it is over three times more effective than other activated carbons and up to five times more effective than synthetic resins (AFT, 2004).

The Carson Cogeneration Plant in California was the first to install the SAG™ system in 1996 to pretreat digester gas. Since then, there have been 47 additional SAG™ systems installations in the U.S. (Tower, 2003b). The installed capital cost of a SAG™ system at the Sanitary District of Decatur (Illinois, US) was about US$ 50,000 (US$ 8.20 per m³/d capacity flow or US$ 0.031 per gpd capacity flow); its annual operating cost is estimated to be approximately US$ 9,500/yr (US$ 1.55 per m³/d capacity flow per yr or US$ 0.006 per gpd capacity flow per yr) (Enochs et al., 2003).

Bergen County used a SAG™ system for pilot testing. Inlet siloxane concentrations ranged from 2 to 4 ppm. The SAG™ system reduced siloxanes to non-detectable limits (Liang and Sheehan, 2002). Similarly, the Sanitary District of Decatur was able to reduce its siloxane levels from between 4ppm-7ppm to below 100ppbv using a SAG™ dual filter system (Enochs et al., 2003). Table 3-2 details other plants that used activated carbon to pretreat digester gas.
Table 3-2. Plants Using Activated Carbon for Preactreating Digester Gas.

<table>
<thead>
<tr>
<th>Plant</th>
<th>Gas flow</th>
<th>Installation</th>
<th>Technology details and efficiencies</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alvardo WWTP, Union City, CA (Slezak et al., 2002)</td>
<td>17000 m³/day (600,300 scf/d)</td>
<td>One unit, 820kg media. Use gas compression, condenser/moisture removal, reheating, and particle filter</td>
<td>Protect Gas Engines</td>
</tr>
<tr>
<td>Annacis Island, Vancouver, BC (Slezak et al., 2002)</td>
<td>45000m³/day (1,589,000 scf/d)</td>
<td>One unit with 500kg media. Treats 800m³ gas/kg media. Use gas compression, condenser/moisture removal, reheating, and particle filter</td>
<td>Protect Gas Engines Treatment involves outlet concentration of 5mg/m³ (survey data)</td>
</tr>
<tr>
<td>Bergen County Utility, Little Ferry, NJ (Tower, 2003b)</td>
<td>8150-32600 m³/day (287,800 - 1,151,000 scf/d)</td>
<td>Implemented full scale: 2 vessels operating series (plus 1 on standby), 3600lb media each (PMG). 3 different types of media in layers</td>
<td>Gas engine and OCR catalyst protection. Inlet 2-4ppm, reduced to non-detectable limits, H2S was also consistently &lt;1ppm in pilot tests (Liang et al, 2002)</td>
</tr>
</tbody>
</table>

Other Techniques for Siloxane Removal: Huppmann et al. (1996), and Glus et al. (1999) have reported successful siloxane removal with several resins. Schweigkofler & Niessner (2001) observed siloxane removal using silica gel on bench and pilot scales.

3.2.3 Hydrogen Sulfide
Fuel gases containing high concentrations (above 10ppm) of sulfur compounds are referred to as sour gases and sweet gases refer to fuels with sulfur content below 10 ppm (CAT, 1997a). Digester gas is most often “sour” as it typically contains between 100 to 10000ppm (or 0.01 to 1%) of H2S (Osinga, 2000).

3.2.3.1 Impact on Energy Recovery Systems and the Environment
The presence of H2S alone, or in combination with water vapour, is detrimental to gas handling devices. Norris (1943) reported serious corrosion and pitting of metal surfaces due to H2S. Also, when water vapour is present during combustion, sulfuric acid (H2SO4) is formed, which is extremely corrosive to the exhaust side of burners, gas lamps and engines (Fulton, 1991; Katehis et al., 2003). Sulfuric acid can be neutralized by the alkalinity present in the engine oil, but unless the oil is changed at up to three times the normal frequency, corrosion of engine components would occur (Fulton, 1991).

Fulton (1991) states, that H2S is particularly harmful to internal combustion engines. However, the presence of H2S also inhibits the removal of siloxanes and halogenated VOCs (CAT, 1997a; Spiegel et al., 1999). Siloxanes are of particular threat to microturbines, while halogenated VOCs are detrimental to phosphoric acid fuel cells (Spiegel et al., 1999).
In addition to the negative effect H$_2$S has on ERS performance, its main combustion product, sulfur dioxide (SO$_2$) reduces ambient air quality significantly (Constant et al., 1989).

### 3.2.3.2 Hydrogen Sulfide Removal Techniques

Technologies developed for removing H$_2$S from natural gas and refinery gas are fairly well established. However, because digester gas has higher levels of H$_2$S than natural gas and refinery gas, these processes have limited application with digester gas (Osinga, 2000).

A summary of the various H$_2$S removal processes is presented in Table 3-3. The following processes either have current applications in the municipal wastewater treatment industry or have the potential for conditioning of digester gas to remove H$_2$S.

**Table 3-3. Summary of the Various H$_2$S Removal Techniques**

<table>
<thead>
<tr>
<th>Techniques</th>
<th>H$_2$S Inlet Concentrations</th>
<th>Level of H$_2$S Treatment</th>
<th>Waste Product</th>
<th>Capital Cost (US$)</th>
<th>Annual Operation &amp; Maintenance Cost (US$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Activated Carbon</td>
<td>Low</td>
<td>&gt;85%</td>
<td>Spent media</td>
<td>10.38/m$^3$/day$^2$ (0.294/scf/d)</td>
<td>0.004/m$^3$/day$^2$ (0.00011/scf/d)</td>
</tr>
<tr>
<td>Iron Salts Addition</td>
<td>&gt;6,000 ppmv</td>
<td>&gt;85%</td>
<td>Iron sulfide</td>
<td>0.05/m$^3$/d (0.0002/gpd) capacity flow (3)</td>
<td>0.35/m$^3$/d (0.0013/gpd) capacity flow (3)</td>
</tr>
<tr>
<td>Iron Sponge</td>
<td>200-5,000 ppmv</td>
<td>90%</td>
<td>Spent iron sponge (containing iron oxide, sulfur, soda, ash, limestone, wood)</td>
<td>81.00/m$^3$/d (2.29/scf/d) digester gas flow</td>
<td>12/m$^3$/d (0.034/scf/d) digester gas flow</td>
</tr>
<tr>
<td>SulfaTreat</td>
<td>&lt;500 ppmv</td>
<td>&gt;85%</td>
<td>Iron disulfide</td>
<td>9.5/m$^3$/d (0.269/scf/d)digester gas flow (4)</td>
<td>17.5/m$^3$/d (0.496/scf/d) digester gas flow (4)</td>
</tr>
<tr>
<td>BINAX</td>
<td>&gt;99%</td>
<td></td>
<td>Water saturated with HS$^-$</td>
<td>187/m$^3$/d (5.295/scf/d) digester gas flow (5)</td>
<td>Not Available</td>
</tr>
<tr>
<td>LO-CAT</td>
<td>30-8,000 ppmv</td>
<td>&gt;98%</td>
<td>Elemental sulfur</td>
<td>Not Available</td>
<td>6.40/m$^3$/d (0.181/scf/d) digester gas flow</td>
</tr>
<tr>
<td>Apollo</td>
<td>200-20,000 ppmv</td>
<td>&gt;99%</td>
<td>Elemental sulfur</td>
<td>Not Available</td>
<td>6,390/yr</td>
</tr>
</tbody>
</table>

(1) Extracted from Earth Tech (Canada) Inc. et al., 2002 unless indicated
(2) Liu, 2005
(3) Dezham et al., 1988
(4) SulfaTreat, 2002 – quote for a two-vessel system
(5) Henrich, 1984 - Price quoted for a 2,680 m$^3$/d digester gas flow BINAX system

**Activated Carbon Adsorption:** Activated carbon adsorption is a common H$_2$S removal method (Schweigkofler and Niessner, 2001; Earth Tech (Canada) Inc. et al., 2002). The activated carbon used is impregnated with an oxidant (Schweigkofler and Niessner, 2001). In the presence of moisture, the impregnated carbon acts as both an adsorbent and an oxidation catalyst, oxidizing the H$_2$S to elemental sulfur, which is deposited on the carbon particles. However, this alternative
is only suitable for small volumes of gas and low H$_2$S concentrations (Earth Tech (Canada) Inc. et al., 2002).

As mentioned in Table 3-3, the removal efficiency of activated carbon is greater than 85%. In one study, potassium-hydroxide impregnated activated carbon was found to remove 98% of the H$_2$S, reducing the sulfur level to less than 10ppb (Spiegel and Preston, 2000).

Activated Carbon Type CJ is one carbon adsorption process that is used to remove H$_2$S from digester gas. It is currently licensed by Barnebey & Sutcliffe Corp. (Earth Tech et al., 2002; Foral and Al-Ubaidi, 1993), and can handle up to (18.1 kg/d) (40 lbs/day) of sulfur (Foral and Al-Ubaidi, 1993). The carbon is impregnated with ferric oxide, which increases the capacity for H$_2$S eight-fold over untreated activated carbon (Foral and Al-Ubaidi, 1993). A shortcoming of the CJ system is that the spent media is not regenerable and in some jurisdictions, local regulations limit the quantity that can be disposed of in landfills (Foral and Al-Ubaidi, 1993).

**Iron Salts Addition:** Using iron salt precipitation of sulfides is another removal technique. Iron salts used are ferric chloride (FeCl$_3$) and ferrous chloride (FeCl$_2$) at a dosage of 3 and 5 mg/L (0.000025 and 0.000042 lb/gal). These are applied to either the WWTP influent or the anaerobic digester influent (Dezham et al., 1988; Earth Tech et al., 2002; Katehis et al., 2003) as shown in Figure 3-10. FeCl$_3$ is slightly more effective than FeCl$_2$, in oxygen free conditions, which are the conditions found in anaerobic digesters. This results in a lower dosage requirement of FeCl$_3$ than FeCl$_2$ required at a WWTP (Katehis et al., 2003). The reaction involving the addition of iron salts is as follows:

$$\text{Fe}^{2+} + 2\text{Fe}^{3+} + 4\text{HS}^- \rightarrow \text{Fe}_3\text{S}_4 + 4\text{H}^+ \quad (3-1)$$

This method removes H$_2$S by reducing concentrations in the dissolved form, so less escapes into the gas in the digestion process. Several forms of iron sulfide could be formed: pyrrhotite, ferric sulfide, smithite, pyrite, and marcasite (U.S. EPA, 1974; Katehis et al., 2003). There are other competing reactions for iron in the digester as well. Therefore, the actual dose needed can only be determined with full-scale tests (Earth Tech et al., 2002). As indicated in the cases presented in Table 3-4, the dosage and the removal effectiveness differs between WWTPs. Generally, levels below 200ppm of H$_2$S can be achieved with iron salts (Chambers and Potter, 2002), however, it should be noted that iron salts give diminishing returns at low sulphide levels (Walton et al., 2003).

An iron salt dosing system was placed into service at San Jose/Santa Clara Water Pollution Control Plant in San Jose (California, U.S.) in February 1986. The chemical addition system, including the storage tank, metering pumps, and piping were constructed at a cost of about US$0.05/m$^3$/d capacity flow (US$0.0002 per gpd capacity flow). According to (Dezham et al., 1988), the annual chemical cost for adding FeCl$_2$ to reduce H$_2$S in digester gas to less than 300 ppmv was US$0.35/m$^3$/d capacity flow (US$0.0013 per gpd capacity flow). However, the cost of iron salt addition is dependent on the local cost of the iron salt, FeCl$_2$ or FeCl$_3$, (Earth Tech et al., 2002).

There are many operational issues to consider when using iron salts. JWPCP used iron salts (FeCl$_2$) to further reduce H$_2$S from 200 ppm to 40 ppm, but found any extra FeCl$_2$ addition
negatively impacted the digestion process: volatile acids increased, and gas production rate and alkalinity decreased (Earth Tech et al., 2002).

![Diagram of wastewater treatment processes with iron salt options]

**Figure 3-10 Schematic Representation of an Iron Salt Process.**

**Table 3-4. Iron Chloride Dosages for H₂S reduction**

<table>
<thead>
<tr>
<th>Plant</th>
<th>H₂S reduction</th>
<th>Dosage</th>
<th>Other factors</th>
</tr>
</thead>
<tbody>
<tr>
<td>City of Santa Maria WWTP, CA</td>
<td>2900 to 950 ppm</td>
<td>16 kg FeCl₂/1000 kg VSS (16 lb FeCl₂/1000 lbVSS)</td>
<td>Further reduction to 300 ppm with 3.9 mg/l (0.0002 lb/scf) FeCl₂ or 4.4 mg/l (0.0003 lb/scf) of FeCl₃ added</td>
</tr>
<tr>
<td>City of Lethbridge WWTP, Alberta</td>
<td>2400 ppm to 600 ppm</td>
<td>600 L/d (21.2 scf/d) 37% FeCl₃</td>
<td>HRT reduced from 40 to 25 days decreased H₂S 6800 ppm to 2400 ppm</td>
</tr>
<tr>
<td>San Jose/Santa Clara WPCP, CA</td>
<td>300 ppm final concentration</td>
<td>5-9.5 kg (11-30 lb) FeCl₂ or 15-27 kg (33-60 lb) FeCl₃</td>
<td></td>
</tr>
<tr>
<td>Encina WPCF, Carlsbad, CA</td>
<td>From &gt;1000 ppm to 200-300 ppm</td>
<td>Continuous chemical feed reduced demand for the ferric compared to batch feeding</td>
<td></td>
</tr>
<tr>
<td>Sacramento, CA (Kido et al, 1995)</td>
<td>1500 ppm to 165 ppm</td>
<td>Added iron chloride to headworks.</td>
<td></td>
</tr>
<tr>
<td>Longmont WWTP, Colorado (Bielefeldt et al, 2002)</td>
<td>52-95% reduction for 1.2 – 74 ppm conc</td>
<td>2 mg/L (0.0001 lb/scf) ferric chloride to headworks (42% solution - ~33 gpd into ~9 mgd flow)</td>
<td></td>
</tr>
</tbody>
</table>

Earth Tech et al., 2002 unless otherwise indicated.
The advantage of iron salt addition includes its low capital cost; ease of process control; and generation of a non-hazardous end product, iron sulfide, (Earth Tech et al., 2002; Katehis et al., 2003). This process can decrease the biogas H2S content to lower than levels required for use in most engine-generators (to about 100-300 ppmv) (Katehis et al., 2003).

The disadvantages of this alternative are that iron salts, with a pH of less than 1, can be corrosive. Also, sludge production increases as a result of iron precipitation, and formation of ferrous phosphate or vivanite (Fe3PO4·8H2O) scaling when iron salts are added directly to the digester feed (Earth Tech et al. 2002; Katehis et al., 2003). To alleviate the problem of vivanite buildup, Dezham et al. (1988) suggested adding iron salts to the plant influent instead; no plugging of lines would occur because precipitants formed in primary sedimentation tanks would be bound in primary sludge and transported to digesters and the effect of iron salts addition to plant influent on subsequent treatment efficiencies for removal of BOD5, TSS, or nitrogen would not be significant (Dezham et al., 1988).

The relative cost of iron salt pretreatment in comparison to other available H2S removal techniques varies between WWTPs. Although iron salts have a low capital cost, the operating cost is high due to chemical use. Therefore, this may not be the most cost-effective method of removing H2S. It has been found to be cost effective for lowering H2S levels that are marginally higher than limit for gas engines, 100-300ppm, for example (Katehis et al., 2003).

Iron Sponge Adsorption: Iron sponge adsorption is an iron oxide-based scavenging process for removal of H2S from digester gas. It is capable of reducing the H2S concentration in digester gas to about 35 ppmv. Sacramento Regional WWTP (California, U.S.), Union Sanitary District WWTP (California, U.S.), Central Marin Sanitation Agency WWTP (California, US), and the JWPCP are some of the WWTPs that are using this alternative to control H2S in their digester gas (Earth Tech et al., 2002; Kido et al., 1995). The system consists of stainless steel, cylindrical vessels with removable covers. The iron sponge media is wood chips or granular activated carbon impregnated with hydrated ferric oxide (2Fe2O3·H2O). A fiberglass grate supports the media (Kido et al., 1995). The chemical reaction of the process is as follows:

\[ 2\text{Fe}_2\text{O}_3 + 6\text{H}_2\text{S} \rightarrow 2\text{Fe}_2\text{S}_3 (s) + 6\text{H}_2\text{O} \]  

A schematic representation of an iron sponge system is presented in Figure 3-11.

Low-pressure gas flows through the dry media, and the H2S in the influent gas stream combines with the iron oxide to form iron sulfide (Fe2S3) and water (Earth Tech et al, 2002; Kido et al., 1995; Ravishanker and Hills, 1984). Gas should be wet when passing through the iron sponge or drying of the bed will cause the bed to reduce its reactive capacity. If the gas is not already water-saturated or if the influent stream has a temperature higher than 50°C (120°F), water with soda ash is sprayed into the contactor to maintain desired moisture and alkaline conditions (Foral and Al-Ubaidi, 1993).
The media can be regenerated by filling the vessel with water and passing oxygen through the bed to convert the iron sulfide back to iron oxide and elemental sulfur (Earth Tech et al., 2002; Kido et al., 1995; Ravishanker and Hills, 1984):

\[ 2\text{Fe}_2\text{S}_3 + 3\text{O}_2 \leftrightarrow 2\text{Fe}_2\text{O}_3 + 6\text{S}_\text{(s)} \]  

(3-3)

Elemental sulfur is non-hazardous and can be applied to a landfill or used as an agricultural product (Earth Tech et al., 2002). However, the spent media cannot be regenerated to its original capacity because it becomes coated with elemental sulfur, which blocks the media surface and increases the pressure drop across the bed. The media can be regenerated two to three times during its lifetime of about three years; however, it only regains 50-60% of its original capacity after regeneration (Kido et al., 1995).

Whessoe Varec Biogas’ 235 Series Gas Purifier has a means of introducing a controlled volume of air to the process to provide continuous regeneration of the iron sponge. (Earth Tech et al., 2002; Henker, 2003). A continuous regeneration system can extend the bed life by four to five times. It provides more efficient H₂S removal and allows for a higher gas flow rate (Earth Tech et al., 2002). In parallel iron sponge operation, the inlet H₂S concentration range is 2500 ppm, but can treat higher concentrations if operated in series. The operating temperature of the Whessoe Varec system is between 2-49°C (36-120°F), which requires a supply of 620-690kPa (90-100psig) compressed air, and water at 275-480 kPa (40-70 psig). Single units have various media sizes to treat flow rates of about 550-4000 m³/day (145,295-1,056,688 gpd). Varec units are reported to reduce H₂S inlet concentrations to less than 4.5 ppm (Varec Biogas, 2002)
The main advantage of the iron sponge is its capability of handling high inlet \( \text{H}_2\text{S} \) concentrations (around 5,000 ppmv) though more than one unit is required for higher concentrations (Henker, 2003). The Iron Sponge licensed by Connelly-GPM Inc. can handle up to 45.36 kg sulphur/d (100lbs sulfur/day) (Foral and Al-Ubaidi, 1993). They are also considered reliable, simple to operate, and have low energy and maintenance requirements.

The predominant disadvantage is that the iron sponge regeneration process can be dangerous, as the scavenging medium can release sulfur in an exothermic reaction. When exposed to air it can result in fire (Earth Tech et al., 2002; Katehis et al., 2003; Osinga, 2000). To prevent spontaneous combustion, the sponge must remain moist at all times (Katehis et al., 2003). Some jurisdictions consider spent iron sponge media hazardous material and require it to remain moist at all times to prevent autoignition. In order to dispose the media as non-hazardous material, it must be sufficiently re-oxidized so that it no longer poses a threat of ignition (Foral and Al-Ubaidi, 1993).

Other disadvantages of iron sponges include their high capital, which is directly related to its installation costs; their high operational costs caused by the high media replacement requirements; and subsequently their large labor requirements.

Iron sponges often are a good option for low biogas volume/low sulfide loading applications (Katehis et al, 2003). Based on an assumed gas flow rate of 3,870 m³/d (1,022,346 gpd) and an inlet \( \text{H}_2\text{S} \) concentration of 1500 ppmv, the installed capital cost of an iron sponge system is approximately US$ 312,500 (US$ 81/m³/d digester gas flow rate or US$ 2.294 scf/d digester gas flow rate), and the annual operating cost is about US$ 46,000 (US$ 12/(m³/d digester gas)/yr or US$ 0.340 /(scf digester gas)/yr) (Earth Tech et al., 2002). Table 3-5 lists some WWTPs that use an iron sponge.

<table>
<thead>
<tr>
<th>Plant</th>
<th>( \text{H}_2\text{S} ) reduction</th>
<th>Dosage</th>
</tr>
</thead>
<tbody>
<tr>
<td>Union Sanitary District WWTP (Earth Tech et al., 2002)</td>
<td>200 to 50 ppm</td>
<td>8 units</td>
</tr>
<tr>
<td>Central Marin Sanitation Agency WWTP, San Rafael, CA (Kido et al, 1995)</td>
<td>400 to 35 ppm</td>
<td>2 units in series (replaced media 3 times in 8 years)</td>
</tr>
</tbody>
</table>

**SulfaTreat:** The SulfaTreat system (Figure 3-12) consists of a proprietary granular media of ferric oxide and triferric oxide supported on an inert surface packed into one or more vertical pressure vessels operated with a downward gas flow (Al-Issa, 2004; SulfaTreat, 2002; Earth Tech et al., 2002). A small inlet separator should be placed just upstream of the SulfaTreat vessel to remove excess liquid from the gas.

Reported benefits of SulfaTreat relative to iron sponge are that it does not ignite spontaneously, removes two to three times more \( \text{H}_2\text{S} \), and last two to three times longer (NATCO Group, 2002). The SulfaTreat system has been used since 1998, mainly in the natural gas and oil industries. There is a lack of operational history of SulfaTreat applications on digester gas (Osinga, 2000). However, the manufacturer states that gas inlet concentrations can be as high as 20,000 – 30,000 ppmv, and are purified to non-detectable limits (Al-Issa, 2004). \( \text{H}_2\text{S} \) removal to less than 1 ppmv in the pretreated gas was reported by Foral and Al-Ubaidi (1993) for water-saturated gas streams at temperatures between 10-50°C (50-122°F).
Operation in cold weather is generally not a problem as the reaction is exothermic and will automatically warm the gas (Al-Issa, 2004). The reaction rate increases substantially with temperature. A six-fold increase was found when the temperature increased from 4-60°C (40-130°F). Operation below 4°C (40°F) is not recommended, and high temperatures can dry out the bed, which decreases the reaction rate (Foral and Al-Ubaidi, 1993).

The process is also not affected by system pressure or the presence of additional contaminants in the gas stream, but saturated gas achieves the best results. Optimum efficiency is achieved by monitoring the water content of the digester gas and injecting water when required to maintain a saturated gas (Osinga, 2000; Earth Tech et al., 2002). The standard system used for digester gas can handle a maximum of 2 psig so digester gas can generally be treated directly from the digester without compression (Al-Issa, 2004).

The system has no mechanical parts and does not require energy to operate (Al-Issa, 2004). Digester gases with high H₂S concentrations would typically be operated in series. The amount of media required (hence the vessel size) would be determined by the H₂S concentration and the desired time between media changing (Earth Tech et al., 2002). For example, two vessels (5.5m bed height, 1.8m diameter or 18.04 ft bed height, 5.91 ft diameter) containing 16300kg (35935 lb) of media would last 79 days between media change outs (Osinga, 2000). About 450g of media/50g H₂S removed (0.992 lb/0.110 lb) is required (Earth Tech et al., 2002) or 1kg (2.2 lb) of media is required per 0.11kg (0.243 lb) of H₂S removed (Osinga, 2000).
The by-product of the Sulfatreat process is iron disulfide, which is stable and can be disposed of as non-hazardous waste. The media is non-regenerable and non-hazardous, however, disposal by landfill depend on local regulations (Earth Tech et al., 2002 and Osinga, 2000). Alternatively, the spent media can be used as construction or agricultural material (Earth Tech et al., 2002).

The Sulfatreat capital and operating costs are relatively low, and operation is relatively simple. Biogas from WWTPs can use Sulfatreat typically for gases with relatively low sulfur concentrations, less than 0.5 tpd, (Earth Tech et al., 2002).

The City of Lakeland WWTP (Florida, US) and the Stickney Water Reclamation Plant, in Chicago, (Illinois, US) are currently using Sulfatreat to polish gases with relatively low sulfur concentrations (Earth Tech et al., 2002). The capital cost of a Sulfatreat system is estimated to be about US$ 9.50/m³/d digester gas flow (US$ 0.27 scf/d digester gas flow) and the associated annual operating cost is approximately US$ 17.50/ (m³/d digester gas flow)/year or US$ 0.50/ (scf/d digester gas flow)/y for an average digester gas flow rate of 3870 m³/d (1,022,000 gpd) with an average H₂S concentration of 1500 ppmv (Earth Tech et al., 2002).

**Water Scrubbers:** Water-only scrubbers operate on the principle of H₂S solubility in water (Osinga, 2000). H₂S absorbed in the water dissociates into HS⁻ and S²⁻ ions. Chemical equilibrium depends on the H₂S concentration, solution pH and temperature (Earth Tech et al., 2002). Under atmospheric pressure, large volumes of water would be required. Less water is required as the working pressure increases. Both CO₂ and H₂S are acidic gases, and effective removal by water scrubbing requires that the water solution pH be alkaline. Caustic soda is commonly used for this purpose. Literature reports that water scrubbing at 13.6 to 20.4 atm (200 – 300 psi) can reduce CO₂ to 4-5% (Constant et al., 1989). A generic water scrubber can take in 200- 2500 ppm H₂S (Earth Tech et al., 2002). Removal of H₂S from off-gases from municipal wastewater treatment are typically very high with packed scrubbing towers, achieving up to 99.5 % removal efficiency (Ceilcote, 2006).

**BINAX:** The BINAX system is a water scrubber that uses high pressure scrubbing to remove both H₂S and CO₂ (details on CO₂ removal are discussed in the next section) (Henrich, 1984; Earth Tech et al., 2002). The reactions that take place in a wet scrubber depend on pressure, temperature, and the pH of the water as they influence the solubility levels of the gases in the aqueous solution (Earth Tech et al., 2002). At neutral pH, the minimum pressure is 60psig for H₂S removal only, and the maximum is 300psig, which is required to remove both H₂S (to less than 1ppm), and CO₂ (Osinga, 2000).

Figure 3-13 illustrates a schematic representation of a BINAX system. The basic BINAX system consists of two towers. One tower purifies the digester gas with pressurized water and the other tower removes the contaminants from the water and allows water to recirculate. To process biogas in the basic BINAX system, the crude digester gas is first compressed and pressurized, then injected into the base of the pressurized tower (gas scrubber tower). As the crude digester gas flows up the first tower, a counter flow of water absorbs the contaminants leaving the purified methane to exit the top. The water from the bottom of the scrubber tower is depressurized and piped to the top of the water regeneration tower. The BINAX uses 3.4 gpm of water per cfm of gas treated. In the regenerator tower the CO₂ and H₂S are flashed from the water and vented to
atmosphere. The water is then pumped back to the top of the gas scrubber tower. Make-up water is required to replace evaporation losses and periodic blow down. Typically H₂S is reduced to 4 ppmv in a BINAX system (Henrich, 1984).

**Figure 3-13. Schematic Representation of a BINAX System.**

The main advantages of this system are its simplicity of design and operation, its ease of maintenance, and the fact that no chemicals are required (Henrich, 1984). The main disadvantage is the high capital cost. The capital cost, including the pump, compressor, scrubber and controls, is US$ 500,000 for a 2,680 m³/d (94,643 gpd) system (US$ 187/m³/d capacity flow or US$ 0.708 per gpd capacity flow); the operating costs are primarily due to the energy to compress the gas and to pressurize the plant water to be used in the operation, which in some cases can be high (Earth Tech et al., 2002). For a 33980 m³/day (1.2M ft³/day) capacity system, water pump and gas compressor power are 250hp and 350hp respectively (Osinga, 2000). No gas flow rate ranges from manufacturer were found. Actual installed and available sizes have been reported between 425-42500 m³/day (15000 - 1.5M scf/d) (Henrich, 1984).

**LO-CAT:** LO-CAT is a patented, wet scrubbing, liquid reduction-oxidation (redox) system that uses a chelated iron solution to convert H₂S to elemental sulfur in slurry form (Gas Technology Products LLC, 2003; Earth Tech et al., 2002). Chelating agents are used to keep the iron from precipitating as it alternates between reduced (ferrous ion) and oxidized (ferric ion) states. Two different redox reactions take place: one in the absorber section, converting H₂S to elemental sulfur, and one in the oxidizer section, which regenerates the catalyst (see Figure 3-14). The absorber reactions are as follows (Gas Technology Products LLC, 2003):

\[
H_2S(g) \leftrightarrow H_2S(aq) \quad (3-4)
\]

\[
H_2S(aq) \leftrightarrow HS^- + H^+ \quad (3-5)
\]

\[
HS^- + 2Fe^{3+} \rightarrow S(s) + 2Fe^{2+} + H^+ \quad (3-6)
\]
In the absorber, H$_2$S is absorbed into the slightly alkaline, aqueous LO-CAT solution. The H$_2$S ionizes to bisulfide, which is oxidized to sulfur by reducing the iron ion from the ferric to the ferrous state. The reduced iron ions are then transferred from the absorber to the oxidizer. In the oxidizer, atmospheric oxygen is absorbed into the LO-CAT solution. The oxidizer section reactions are as follows (Gas Technology Products LLC, 2003):

$$O_2(g) \leftrightarrow O_2(aq) \quad (3-7)$$

$$O_2(aq) + 4Fe^{2+} + 2H_2O \rightarrow 4Fe^{3+} + 4OH^- \quad (3-8)$$

The ferrous iron is reoxidized into ferric iron, which regenerates the catalyst. The regenerated catalyst is ready for use in the absorber section. The overall reaction is a modified Claus reaction as follows (Gas Technology Products LLC, 2003):

$$H_2S(g) + \frac{1}{2}O_2(g) \rightarrow H_2O + S^0 \quad (3-9)$$

Several types of chemicals need to be added to the system to maintain the above reactions at a high pH, to replace chelated iron lost in the sulfur removal process, and to replace the degraded chelating agents. The spent catalyst solution is regenerated (oxidized) with air, although some catalyst is lost in the slurry (Gas Technology Products LLC, 2003). The process does not use any toxic chemicals or produce any hazardous waste. The reaction is exothermic and may be used in cold climates. The blower for the oxidizing air adds about 50-70°C (122-158°F) to the ambient air temperature (thereby heating the inlet air) (Gas Technology Products, 2003).
The system involves a number of energy spending components, such as pumps, blowers, and solution coolers. The major operating costs are due to chemical costs and electricity and depend on the amount of H2S removed, but the power cost tends to be constant when the unit is running (Gas Technology Products LLC, 2003).

Overall, the system is relatively complex and requires skilled operators (Katehis et al., 2003; Earth Tech et al., 2002) and is designed for large WWTPs (Gas Technology Products LLC, 2003; Katehis et al., 2003). The City of Los Angeles’ Hyperion WWTP (California, US) has reported chemical consumption costs of US$ 25,000/yr or US$ 6.40/(m³/d digester gas flow)/year or US$ 0.18/(scf/d digester gas flow)/y (Earth Tech et al., 2002).

The manufacturer claims removal efficiencies can be greater than 99.9% by design (Gas Technology Products, 2003). When the inlet sulfide concentrations are about 500 ppmv, the effluent concentration is below 30 ppmv with one LO-CAT unit and below 1 ppmv with two units of LO-CAT.

There are several WWTPs that use LO-CAT to control H2S, which are outlined in Table 3-6 (Gas Technology Products LLC, 2003; Katehis et al., 2003; Earth Tech et al., 2002).

<table>
<thead>
<tr>
<th>Plant</th>
<th>H2S reduction</th>
<th>Flow</th>
<th>Other factors</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quote for Earth Tech</td>
<td>1500avg (6000ppm max) to 200ppm</td>
<td>4000 m³/day (1.6 MGD)</td>
<td>Saves at least $1M compared to iron salt addition to achieve 40ppm H2S</td>
</tr>
<tr>
<td>Hyperion, Los Angeles, CA</td>
<td>500ppm down to 30ppm with 1 unit, 1ppm with 2.</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Universal foods, Baltimore, MA</td>
<td>30 000ppm to 850ppm and after modification: 6000ppm to 600ppm</td>
<td>20400m³/day (5.39 MGD)</td>
<td>Had some problems with sulfur build-up and carbonate scaling</td>
</tr>
<tr>
<td>South Bend WWTP, Winnipeg</td>
<td>99% removal for 3000ppm influent</td>
<td>5000m³/day (1.32 MGD)</td>
<td></td>
</tr>
<tr>
<td>Ellsemere Port, England</td>
<td>500 to 8000ppm inlet. Target is 500ppm</td>
<td>6000m³/day (1.59 MGD)</td>
<td>2-stage system, operated intermittently</td>
</tr>
<tr>
<td>Berrihill, England</td>
<td>5-6000ppm H2S</td>
<td>5400m³/day (1.43 MGD)</td>
<td>2 stage, operated intermittently, energy consumption is high</td>
</tr>
</tbody>
</table>

Earth Tech et al., 2002

Apollo: The Apollo gas scrubber removes sulfides and particulates from anaerobic digester gas that contains H2S concentrations up to 20,000 ppmv, at a flow rate of 3,900 to 193,000m³/d (100 to 5000 scfm) (OCETA, 2002a; Earth Tech et al., 2002). The scrubbing process utilizes a modified flotation cell for the absorption of up to 99% of the H2S from the gas stream (Earth Tech (Canada) Inc et al. 2002). The absorption process is facilitated by a catalyst and a patented high mass transfer gas-liquid contactor (OCETA, 2002a). The catalyst is iron with ethylenediaminetetraacetic acid (EDTA) as the chelating agent. The chemical reaction is the six step process which is presented as follows (Earth Tech et al., 2002):
Adsorption

\[ H_2S_{(g)} \leftrightarrow H_2S_{(aq)} \] \hspace{1cm} (3-10)

Ionization

\[ H_2S_{(aq)} + OH^- \leftrightarrow H_2O + HS^- \] \hspace{1cm} (3-11)

Substitution

\[ HS^- + FeOHEDTA_{(aq)}^{2-} \leftrightarrow FeHSEDTA_{(aq)}^{2-} + OH^- \] \hspace{1cm} (3-12)

Oxidation

\[ FeHSEDTA_{(aq)}^{2-} + FeOHEDTA_{(aq)}^{2-} \leftrightarrow 2FeEDTA_{(aq)}^{2-} + S^0 + H_2O \] \hspace{1cm} (3-13)

Regeneration

\[ H_2O + \frac{1}{2} O_2_{(aq)} + 2FeEDTA_{(aq)}^{2-} \leftrightarrow 2FeOHEDTA_{(aq)}^{2-} \] \hspace{1cm} (3-14)

Overall Reaction

\[ H_2S + \frac{1}{2} O_2 \leftrightarrow H_2O + S^0 \] \hspace{1cm} (3-15)

The system is not affected by fluctuating flowrates or H$_2$S concentrations. Self-induced gas flow in the system can reduce or eliminate the need for a fan, blower or compressor. One type of process unit has a powerful mixing unit to transfer the gas into the scrubbing medium up to 20 times faster than conventional liquid contactors (OCETA, 2002b).

A schematic representation of the Apollo system is presented in Figure 3-15, which consists of a scrubber unit vessel and regeneration system vessel (OCETA, 2002a; Earth Tech et al., 2002). The regeneration happens in the vessel, which uses atmospheric oxygen to convert H$_2$S from the scrubbing solution to elemental sulfur. The elemental sulfur is usually fed to a sulfur slurry storage tank and periodically mixed with the digester biosolids and dewatered for final disposal or reuse as agricultural product (OCETA, 2002a; Earth Tech et al., 2002).

The Apollo system was first tested at the Metropolitan Toronto Main Treatment Plant (Ontario, Canada), in 1995 (Ontario MOE, 2003; Earth Tech et al., 2002; Katehis et al., 2003). With iron chelate concentrations of 0.25 to 10 g Fe/L (0.0021 – 0.084 lb/gal) and gas flow rates between 4,320 m$^3$/d (1.14 Mgd) and 192,960 m$^3$/d (51 Mgd), tests showed a H$_2$S removal efficiency of 98% for 100-1000ppmv H$_2$S (Earth Tech et al., 2002).
3.2.4 Carbon Dioxide

CO$_2$ is the major non-methane constituent found in digester gas. CO$_2$ in digester gas remains unchanged after combustion. It is discharged from the ERS with the combustion products. CO$_2$ has no heating value so does not contribute to the overall energy content of the digester gas.

In general, CO$_2$ has no impact on the operation of ERSs and is often not removed from the digester gas. As such, the importance of removing CO$_2$ from digester gas is considered less critical than for the other components discussed above, and so less attention has been applied in this report to describing CO$_2$ removal processes.

In internal combustion engines, the presence of CO$_2$ reduces the burning velocity of the digester gas (Bari, 1996), and the peak pressure inside the engine cylinders (Jawurek et al., 1987). This results in a reduced methane combustion efficiency, which decreases the maximum power output of the engine (Bari, 1996; Jawurek et al., 1987). Jawurek et al. (1987) found that engines operating on digester gas containing more than 30% CO$_2$ were particularly susceptible to harsh and irregular running, which was alleviated when gasoline was supplied to the internal combustion engine simultaneously with the digester gas. The survey of WWTPs found most surveyed plants use natural gas to supplement digester gas with high CO$_2$ levels instead of gasoline.

Water and caustic scrubbers are the simplest type of CO$_2$ removal system. The dilute caustic solution is typically discharged downward through trays or a packing material, while the CO$_2$ rich gas stream is fed through the tower in co-current or counter-current flow.

The BINAX system is one identified operating process for removing CO$_2$ from digester gas. It is able to reduce CO$_2$ from 45% to 2% when high pressure is used. The Renton South WWTP (Washington, U.S.) used BINAX for pipeline injection and the Plantation WWTP (North Carolina, US) used it to produce fuel for onsite vehicles. The Plantation WWTP has since stopped
using BINAX, which had been installed in 1998. They had used the gas in modified 6 cylinder gasoline onsite vehicles. The vehicles were reportedly overheating, and had problems with the gas spraying into the carburetor. To combat this problem, frequent maintenance was performed on the compressor. The Plantation WWTP also had to replace the original carbon steel and cast iron containers with stainless steel ones, as the originals had corrosion problems (Sullivan, 2004).

### 3.3 Digester Gas Energy Recovery Systems

#### 3.3.1 Boilers

Boilers have been in use for energy (heat) recovery from digester gas the longest of any of the ERSs at POTWs. Boilers are robust recovery devices, requiring little in the way of gas pretreatment. Digester gas is combusted to produce hot water or steam. The recovered heat can be used to maintain the anaerobic digester temperature using heat exchangers or direct steam injection, and also for space heating. Boilers have a good energy recovery efficiency, ranging on the order of 75-80 %. The reported range of capital cost for boilers is US$28/MJ (US$23,500/bhp) (Burrowes, 2000), while the annual operating cost is about US$0.56/MJ (US$471/bhp) (Earth Tech (Canada) Inc. et al., 2002).

The energy recovery from digester gas using a boiler is depicted schematically in Figure 3-16.

![Figure 3-16. Schematic of Energy Recovery from Digester Gas using Boilers.](image)

#### 3.3.2 Engine Generators

Digester gas has been used for many years in internal combustion engines (IC engines) that drive electricity generators. When heat is recovered from the engine/generator set, the process is called cogeneration. Companies such as Waukesha, Cooper, Caterpillar, and Jenbacher
manufacture gas engines for cogeneration running on digester gas (Chiu, 2004). The amount of fuel energy recovered as useable thermal energy ranges from 45-50% and the amount of fuel energy converted to electricity ranges between 30% and 35% (Burrowes, 2000; Earth Tech (Canada) Inc et al., 2002; Osinga, 2000). Contaminants in the digester that must be reduced prior to combustion include moisture, H₂S and siloxanes, a component of many cosmetic preparations. The total capital cost of gas engines manufactured by Waukesha is approximately between US$1000/kW and US$1250/kW (Earth Tech et al., 2002). Bautista (1999) reported capital costs varying between $500/kW and $1,400/kW for equipment sizes between 20 kW and 20 MW and efficiencies between 28% and 45%. For the same ranges of equipment size and efficiency O&M costs are between $0.007/kWh and $0.020/kWh (Bautista, 1999). Wartsila (2002) reported O&M costs between $0.0009/kWh (for a capacity of 5,000 kW) and $0.002/kWh (for a capacity of 100 kW). Because the engines combust digester gas with air, the engines may emit elevated concentrations of CO₂ and NOₓ.

A schematic diagram of the ERS of an engine cogeneration set is depicted in Figure 3-17.

![Figure 3-17. Schematic of Energy Recovery from Digester Gas using Engine Generators.](image)

### 3.3.3 Microturbines

Microturbines are a newer ERS that can be used in designated ozone Non-Attainment Areas because of the low NOₓ emissions produced during combustion of the digester gas. The combusted gas drives turbine fan blade on the shaft, which rotates through the generator section. The main providers of microturbine operating on digester gas are Capstone and Ingersoll-Rand. The microturbine generates variable frequency (50/60 Hz) 3-phase AC power. Systems can be upgraded to switch automatically between grid and stand-alone operation. The efficiency of the microturbines is approximately 26-27% at 30 kW operation (Capstone), although both power and efficiency decline above 18°C (65 °F). With combined heat and power (CHP) recovery, the overall efficiency can rise to the order of 70-90%, according to manufacturer’s literature. Removal of
water vapor and siloxanes is recommended. The capital and annual operating costs are estimated to be US$1,450/kW and US$130/kW/yr (about $0.015/kWh), respectively (Earth Tech (Canada) Inc et al, 2002). According to Bautista (1999) Microturbines with capacities between 30 kW and 300 kW have a capital cost varying between $600/kW and $1000/kW and an O&M cost between $0.003/kWh and 0.01/kWh.

A schematic of a microturbine unit is presented in Figure 3-18. Identified microturbine installations at POTWs are summarized in Table 3-7.

![Figure 3-18. Schematic Diagram of Energy Recovery from Digester Gas from a Microturbine Unit.](image)

<table>
<thead>
<tr>
<th>Plant</th>
<th>Location</th>
<th>Date Started</th>
<th># Unit(s)</th>
<th>Installation Capacity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Jeannette</td>
<td>Jeannette, Pennsylvania</td>
<td>2001</td>
<td>1</td>
<td>30kW</td>
</tr>
<tr>
<td>Allentown</td>
<td>Allentown, Pennsylvania</td>
<td>2001</td>
<td>12</td>
<td>360kW</td>
</tr>
<tr>
<td>Colorado Springs</td>
<td>Colorado Springs, Colorado</td>
<td>2001</td>
<td>2</td>
<td>60kW</td>
</tr>
<tr>
<td>San Elijo</td>
<td>Cardiff, California</td>
<td>2001</td>
<td>3</td>
<td>90kW</td>
</tr>
<tr>
<td>Inland Empire Regional Plant-1</td>
<td>California</td>
<td>2001</td>
<td>8</td>
<td>240kW</td>
</tr>
<tr>
<td>Temecula</td>
<td>California</td>
<td>2001</td>
<td>2</td>
<td>60kW</td>
</tr>
<tr>
<td>Chiquita</td>
<td>Santa Margarita, California</td>
<td>2001</td>
<td>2</td>
<td>60kW</td>
</tr>
<tr>
<td>Escondido</td>
<td>California</td>
<td>2001</td>
<td>12</td>
<td>360kW</td>
</tr>
<tr>
<td>Lewiston</td>
<td>Lewiston, New York</td>
<td>2001</td>
<td>2</td>
<td>60kW</td>
</tr>
<tr>
<td>Essex Junction</td>
<td>Essex Junction, Vermont</td>
<td>To be operational</td>
<td>N/A</td>
<td>N/A</td>
</tr>
<tr>
<td>Owl's Head</td>
<td>New York City, New York</td>
<td>To be operational</td>
<td>1</td>
<td>30kW</td>
</tr>
</tbody>
</table>

### 3.3.4 Fuel Cells

Fuel cells operate by converting a hydrocarbon fuel and gas phase oxidant over an electrolyte to produce electricity and heat. Fuel cells can be operated in environmentally sensitive areas because they produce very low emission rates of NOx and SOx, as well as low emissions of CO2. A number of different types of fuel cells are available, operating at different temperatures and using different electrolytes, as indicated in Table 3-8 (Chiu, 2004).
Table 3-8. Types of Fuel Cells.

<table>
<thead>
<tr>
<th>Classification</th>
<th>Type</th>
<th>Abbreviation</th>
<th>Operating Temperature(^1)</th>
<th>Characteristics/ Uses(^2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Low Temperature Fuel Cells (LTFCs)</td>
<td>Phosphoric Acid</td>
<td>PAFC</td>
<td>180°-210°C (356-410°F)</td>
<td>Medium cogeneration systems, steam generation</td>
</tr>
<tr>
<td></td>
<td>Proton Exchange Membrane</td>
<td>PEMFC</td>
<td>80°-90°C (176-194°F)</td>
<td>Used in automobile</td>
</tr>
<tr>
<td></td>
<td>Alkaline</td>
<td>AFC</td>
<td>~80° (176°F)</td>
<td>Used in space vehicles</td>
</tr>
<tr>
<td>High Temperature Fuel Cells (HTFCs)</td>
<td>Molten Carbonate</td>
<td>MCFC</td>
<td>600°-700°C (1110-1290°F)</td>
<td>Large cogeneration systems, range of fuels</td>
</tr>
<tr>
<td></td>
<td>Solid Oxide</td>
<td>SOFC</td>
<td>800°-1000°C (1470-1830°F)</td>
<td>All size of cogeneration systems</td>
</tr>
</tbody>
</table>

\(^1\)Source: Appleby, 1996
\(^2\)Source: Hordeski, 2003

Fuel cells can be contaminated easily by a number of contaminants. Table 3-9 summarizes the digester gas contaminant limits for fuel cell applications.

Table 3-9. Digester Gas Contaminant Limits for Fuel Cell Applications.

<table>
<thead>
<tr>
<th>Digester Gas Contaminants</th>
<th>Concentration limits(^1)</th>
<th>Issue/concern(^1)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen sulfide</td>
<td>&lt; 4ppmv(^2)</td>
<td>Poison fuel processor reforming catalyst</td>
</tr>
<tr>
<td>Halogens (F, Cl, Br)</td>
<td>&lt; 4ppmv</td>
<td>Corrosion of fuel processor components</td>
</tr>
<tr>
<td>Non-Methane Olefinic Carbons</td>
<td>&lt; 0.5% olefins</td>
<td>Poison to fuel processor shift catalysts</td>
</tr>
<tr>
<td>Oxygen</td>
<td>&lt; 0.5%</td>
<td>Over-temperature of fuel processor beds due to excessive oxidation</td>
</tr>
<tr>
<td>Ammonia</td>
<td>&lt; 1ppmv</td>
<td>Fuel cell stack performance</td>
</tr>
<tr>
<td>Water</td>
<td>Remove</td>
<td>Damage to fuel control valves. Transport of bacterial phosphates</td>
</tr>
<tr>
<td>Bacteria/solids</td>
<td>Remove</td>
<td>Fouling of fuel processor piping/beds</td>
</tr>
</tbody>
</table>

\(^1\) Source: Spiegel and Preston, 1999; Spiegel and Preston, 2000; unless otherwise indicated, \(^2\)Masemore and Piccôt (1998) reported a limit for total sulphur < 3 ppmv while Earth Tech et al. (2002) indicated a concentration of H2S less than 0.1 ppmv.

The Project Team identified three main companies using fuel cell technology powered by anaerobic digester biogas. The manufacturers are Fuel Cell Energy Inc., Fuel Cell Technologies Ltd., and UTC Fuel Cells (Table 3-10). Each of these manufacturers uses a different technology for their fuel cells. Fuel Cell Energy Inc. with installations at a King County Seattle facility (Washington, US) and at Los Angeles’ Terminal Island facility (California, US), uses molten carbonate technology for their stationary fuel cells, with power outputs ranging from 250 kW to 2000 kW. Energy recovery efficiency ranges from 47 to 50%. UTC Fuel Cells uses phosphoric acid technology in their product. UTC Fuel Cells has made more significant inroads with installations at Portland’s Columbia Boulevard facility and the Las Virgenes facility in Calabasas (California, US), as well as several sites at New York City POTWs operating with their 200 kW models. Fuel cell installations at the New York treatment facilities are in partnership with the New York Power Authority and the New York State Energy Research and Development Authority (NYSERDA). The reported energy recovery efficiency (electrical and recoverable thermal energy) for UTC Fuel Cells’ phosphoric acid technology is 87%. Fuel Cell Technologies Ltd., which uses solid oxide fuel cell technology, offers modular units of 50 kW, but has no documented installations at this time. The reported energy recovery efficiency of the solid oxide fuel cell from Fuel Cell Technologies Ltd. is 87% considering both electricity and heat recovery. The installations of fuel cell applications at POTWs are summarized in Table 3-11 while operating and technical information for the fuel cell types are provided in Table 3-12.
The PAFC unit capital cost quoted by UTC Fuel Cell is approximately US$ 6,250/kW and its annual operating cost is estimated to be US$ 250/kW/yr (Frankhauser, 2002). This is equivalent to $0.0285/kWh. Bautista (1999) reported a much lower O&M cost, $0.005/kWh to $0.01/kWh for capacities between 5 kW and 3 MW. The author however did not indicate the specific type of fuel cell. A schematic diagram of a fuel cell operating on digester gas is presented in Figure 3-19.

![Figure 3-19. General Schematic of Fuel Cell Cogeneration System.](image)

### 3.3.5 Direct Drive

The energy in digester gas, besides recovered as heat and electricity, can also be recovered for mechanical use. The main applications appear to be for driving air blowers and pumps. Plants identified as using direct drive energy recovery from digester gas are provided in Table 3-13.
### Table 3-10. List of Suppliers of Fuel Cells for Wastewater Treatment Plants.

<table>
<thead>
<tr>
<th>Name of the Fuel Cell Manufacturer</th>
<th>Product(s)</th>
<th>Website</th>
<th>Name of the contact person</th>
<th>Last Name</th>
<th>First Name</th>
<th>Title</th>
<th>Phone</th>
<th>Fax</th>
<th>Email</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>TESTING</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>DFC</td>
<td><a href="http://www.foe.com">http://www.foe.com</a></td>
<td>Baker</td>
<td>Bill</td>
<td>Contacts for News Media</td>
<td>(860)350-9100</td>
<td>N/A</td>
<td>N/A</td>
<td></td>
</tr>
<tr>
<td><strong>COMMERICALLY AVAILABLE</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>PC25 (200kW)</td>
<td><a href="http://www.utcfuelcells.com">http://www.utcfuelcells.com</a></td>
<td>Frankhauser</td>
<td>Greg</td>
<td>Regional Sales Manager</td>
<td>(410)740-5616</td>
<td>N/A</td>
<td></td>
<td><a href="mailto:Greg.Frankhauser@UTCFuelCells.com">Greg.Frankhauser@UTCFuelCells.com</a></td>
</tr>
<tr>
<td><strong>PLANNED FUTURE PROGRAM</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fuel Cell Technologies LTD.</td>
<td>Modular 50kW commercial unit</td>
<td><a href="http://www.fct.ca">http://www.fct.ca</a></td>
<td>Allen</td>
<td>Gary</td>
<td>Sales (SOFC)</td>
<td>(613)544-8222</td>
<td>(613)544-5150</td>
<td><a href="mailto:gallen@fct.ca">gallen@fct.ca</a></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Adams</td>
<td>Mike</td>
<td>Sales (Al/O₂)</td>
<td></td>
<td></td>
<td><a href="mailto:madams@fct.ca">madams@fct.ca</a></td>
<td></td>
</tr>
</tbody>
</table>
Table 3-11. List of Fuel Cell Installations.

<table>
<thead>
<tr>
<th>Fuel Cell Manufacturer</th>
<th>Project Partner</th>
<th>Fuel Cell</th>
<th>Location</th>
<th>Building</th>
<th>Status</th>
<th>Fuel used</th>
<th>Cost of Project(USD)</th>
<th>Comments</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>King County, Washington</td>
<td>1MW DFC®</td>
<td>King County, Washington</td>
<td>Municipal WWTP</td>
<td>Construction-current (April 2003); Operation(October, 2003 expected)</td>
<td>Digester Gas</td>
<td>$1.8-2.2 million USD depending on source</td>
<td>Project follows after LADWP’s 250kW DFC300® in Los Angeles (March 2003)</td>
</tr>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>LADWP ¹</td>
<td>250kW DFC300®</td>
<td>Los Angeles, California</td>
<td>Fuel Cell Power Plant at Terminal Island in San Pedro</td>
<td>To be delivered</td>
<td>Digester Gas from a nearby WWTP</td>
<td>Unknown</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>City of Portland, Oregon</td>
<td>200kW PAFC(PC25)</td>
<td>Portland, Oregon</td>
<td>Columbia Blvd. WWTP</td>
<td>In Operation since May 1999</td>
<td>Digester Gas</td>
<td>~$1.3 million USD</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>Energy 2000 Las Virgenes Municipal Water District and Truunfo Sanitation District</td>
<td>Two 200kW PAFC (PC25)</td>
<td>Calabasas, CA</td>
<td>Rancho Las Virgenes Composting Facility</td>
<td>In Operation since 1999</td>
<td>Digester Gas from nearby WWTP</td>
<td>~$2.6 million USD</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>NYPA, KeySpan With Westchester County WWTP in Yonkers, NY</td>
<td>200kW PAFC PC25</td>
<td>Yonkers, NY</td>
<td>Westchester County WWTP</td>
<td>In Operation since April 1997</td>
<td>Digester Gas</td>
<td>Supplies Supplemental Power, Grid Parallel, World’s First ADG fueled fuel cell</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>NYPA and NYSERDA ³</td>
<td>Three 200kW PAFC(PC25)</td>
<td>Coster Street and Ryawa Ave, Bronx, NY 10474</td>
<td>Hunts Point WWTP</td>
<td>Under Construction (2003)</td>
<td>Digester Gas</td>
<td>N/A</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>NYPA and NYSERDA</td>
<td>Two 200kW PAFC (PC25)</td>
<td>63 Flushing Ave, Brooklyn, NY 11205</td>
<td>Red Hook WWTP</td>
<td>Under Construction (2003)</td>
<td>Digester Gas</td>
<td>N/A</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>NYPA and NYSERDA</td>
<td>Two 200kW PAFC(PC25)</td>
<td>43-01 Berrian Blvd., Astoria, NY 11105</td>
<td>Bowery Bay WWTP</td>
<td>Under Construction (2003)</td>
<td>Digester Gas</td>
<td>N/A</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>NYPA and NYSERDA</td>
<td>One 200kW PAFC(PC25)</td>
<td>751 Mill Rd., Staten Island, NY 10306</td>
<td>Oakwood Beach WWTP</td>
<td>Under Construction (2003)</td>
<td>Digester Gas</td>
<td>N/A</td>
<td></td>
</tr>
</tbody>
</table>

¹ LADWP: Los Angeles Department of Water and Power
² NYPA: New York Power Authority
³ NYSERDA: New York State Energy Research and Development Authority
<table>
<thead>
<tr>
<th>Company Name</th>
<th>Product</th>
<th>Types</th>
<th>Market Entry Date</th>
<th>Fuel Used</th>
<th>Operating Temperature</th>
<th>Power Output</th>
<th>Efficiency</th>
<th>Waste Heat/Emission</th>
<th>Comment</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>DFC 300 Stationary Fuel Cell Unit</td>
<td>Molten Carbonate</td>
<td>Testing</td>
<td>Fuel Flexible</td>
<td>250kW</td>
<td>47%</td>
<td>From typical natural gas: NOx&lt;0.3ppmv SOx&lt;0.1ppmv CO&lt;10ppmv Exhaust Temperature: 650° F Exhaust Heat Available: 300,000 BTU/hr</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>DFC 3000 Stationary Fuel Cell Unit</td>
<td>Molten Carbonate</td>
<td>Testing</td>
<td>Fuel Flexible</td>
<td>2000kW</td>
<td>50%</td>
<td>From typical natural gas: NOx&lt;0.3ppmv SOx&lt;0.01ppmv CO&lt;10ppmv Exhaust Temperature: ~650° F Exhaust Heat Available: 2.8 mm BTU/hr</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fuel Cell Energy Inc.</td>
<td>DFC 1500 Stationary Fuel Cell Unit</td>
<td>Molten Carbonate</td>
<td>Testing</td>
<td>Fuel Flexible</td>
<td>1000kW</td>
<td>49%</td>
<td>From typical natural gas: NOx&lt;0.3ppmv SOx&lt;0.01ppmv CO&lt;10ppmv Exhaust Temperature: ~650° F Exhaust Heat Available: 1.4 mm BTU/hr</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fuel Cell Technologies LTD.</td>
<td>Modular 50kW Commercial Unit</td>
<td>Tubular SOFC</td>
<td>Planned Future Program</td>
<td>Fuel Flexible for first generation. Second generation will use diesel fuel and furnace oil</td>
<td>700 to 1000°C (1290-1830)</td>
<td>5kW</td>
<td>~90% with co-generation</td>
<td>6kW heat Co-generation system to provide electricity and heat to commercial units using natural gas, propane, bio-gas directly, as well as diesel fuel with a pre-reformer</td>
<td></td>
</tr>
<tr>
<td>UTC Fuel Cells</td>
<td>PC25 200kW PAFC Systems</td>
<td>PAFC</td>
<td>Available</td>
<td>Natural Gas, Digester Gas or both</td>
<td>200kW</td>
<td>87% with co-generation</td>
<td>On 15% O2 dry basis: NOx&lt;1ppmv SOx&lt;negligible CO&lt;2ppmv</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Table 3-13. List of Wastewater Treatment Plants Using Direct Drive Cogeneration System.

<table>
<thead>
<tr>
<th>Plant</th>
<th>City</th>
<th>Type of Energy Recovery System</th>
</tr>
</thead>
<tbody>
<tr>
<td>Valley Creek</td>
<td>Birmingham, AL</td>
<td>Engine Blowers</td>
</tr>
<tr>
<td>North Regional</td>
<td>Dayton, OH</td>
<td>Engine Generators and Engine Pump</td>
</tr>
<tr>
<td>Northside</td>
<td>Durham, NC</td>
<td>Engine Blowers</td>
</tr>
<tr>
<td>Village Creek</td>
<td>Forth Worth, TX</td>
<td>Engine Generators and Engine Blowes</td>
</tr>
<tr>
<td>Nansemond</td>
<td>Virginia Beach, VA</td>
<td>Boilers and Engine Blowers</td>
</tr>
<tr>
<td>Muddy Creek</td>
<td>Winston-Salem, NC</td>
<td>Boilers and Engine Blowers</td>
</tr>
</tbody>
</table>

The thermal and electrical energy recovery efficiencies for direct drive applications are 46% and 26-28%, respectively (Wander Associates, 1993). The capital cost is approximately US$21.70/m³/d (US$82,000/MGD) plant capacity flow, and the annual operating cost is estimated to be about US$1/(m³/d)/yr (US$4,200/MGD/yr) [Note: operating costs were expressed on a different basis in this report and could not be directly compared to other units of expression.] (Wander Associates, 1993). A schematic diagram of energy recovery using direct drive equipment is depicted in Figure 3-20.

Use of digester gas to power direct drive prime movers such as blowers and pumps is of interest to this project because of the reported efficiency for converting the energy in biogas to mechanical energy. Electrically powered pumps and blowers must first take the energy in digester gas, convert it to electricity, involving an efficiency loss, and then convert the electrical energy to mechanical energy, involving another energy efficiency loss. Direct drive units are more efficient because they remove a step from the energy use loop. Digester gas is converted directly to mechanical energy, without the intermediate step of conversion to electricity. The overall energy efficiency is thus higher than for electrically driven units. A number of treatment plant sited in the U.S. Mid-West have been visited to see direct drive units in operation.
Computer searching of the technical literature revealed very little about the operation or economics of direct drive units. The most informative citation reviewed was a report prepared by Battelle Pacific Northwest Labs. This report was prepared in 1981 for the U.S. Department of Energy. The objectives of the study were to: 1) to estimate the energy potential from the anaerobic digestion of municipal wastewater solids, and 2) to assess present (i.e. 1981) gas utilization schemes with respect to energy and cost effectiveness. Three onsite ERS were assessed, including heat generation, mechanical energy generation, and electrical energy generation. Of the three digester gas utilization alternatives, the most energy efficient and cost effective was found to be the direct production of mechanical energy to operate aerators and pumps. The report suggested that between 50 and 80% of the mechanical energy required for wastewater treatment could be provided by digester gas. Waste heat recovered from the engines to aid in heating the digesters and buildings further improves the overall energy efficiency. Total energy costs for the mechanical energy generation were estimated at approximately $0.008/m³ ($30/million gallons) treated, compared to $0.009/m³ ($34/million gallons) and $0.012/m³ ($45/million gallons) for the electrical energy generation and heat generation schemes, respectively.

### 3.3.6 Stirling Engine

Stirling engines operating on digester gas are of interest mainly because they can operate without costly GPS for H₂S or siloxane removal. A Stirling engine is a closed-cycle, regenerative heat engine that uses an external combustion process, heat exchangers, pistons, a “regenerator,” and a gaseous working fluid (typically helium or hydrogen) contained within the engine to convert heat to mechanical work. When the working gas is heated, its pressure increases, pushing the piston, and forcing the hot gas into the cool cylinder. As the gas cools, its pressure decreases, allowing atmospheric pressure to push the piston back down. This process repeats rapidly, and the mechanical work performed can be converted to usable energy. The pistons are joined to a swash plate which converts the linear motion of the pistons to rotate a shaft and drive a generator. Engine cooling is done with water, which transfers naturally with heat exchangers for heat recovery.

The two main ways to raise the power output of a Stirling cycle are to increase the pressure in the first phase by increasing the temperature or to cool the gas in the third phase to a lower temperature. Due to heat transfer and mechanical issues, most Stirling engines currently available have capacities under 5 kW. (Scott et al., 2003) although other report suggests units of 25-200 kW are available (Earth Tech et al., 2002).

The two main types of Stirling engines are kinematic and free piston. The kinematic engine has pistons attached to a drive mechanism that converts the linear motion of the pistons to a rotary motion. Because they have a crankshaft and flywheel, kinematic engines may replace internal combustion engines to provide shaft power. The free piston engine uses harmonic motion mechanics and usually planar springs. The pistons are mounted in flexures and oscillate freely, without any contact, and therefore without any wear. They can be configured to provide whatever voltage and frequency are required. (Scott et al., 2003). Major advantages and disadvantages of the Stirling engine are presented in Table 3-14.
Table 3-14. Advantages and Disadvantages of Stirling Engine System.

<table>
<thead>
<tr>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quiet operation</td>
<td>a long startup time</td>
</tr>
<tr>
<td>a physically small size</td>
<td>cannot quickly change power output</td>
</tr>
<tr>
<td>high reliability</td>
<td>high capital costs</td>
</tr>
<tr>
<td>long life</td>
<td>larger than internal combustion engines with similar power output.</td>
</tr>
<tr>
<td>maintenance-free operation</td>
<td></td>
</tr>
<tr>
<td>reduced emissions</td>
<td></td>
</tr>
<tr>
<td>fuel flexibility</td>
<td></td>
</tr>
<tr>
<td>recovery of waste heat created by other processes.</td>
<td></td>
</tr>
</tbody>
</table>

Stirling engines can reduce NOx emissions by 75% or more, compared to reciprocating engines. The technology promises much lower maintenance cost as oil changes are not required, and the internal engine components are not exposed to either H2S or siloxanes. Currently the capital cost of the Stirling engine is still over 60% higher than IC engines but these costs are predicted to drop dramatically with time and increased production of the units (Chambers and Potter, 2002).

One company, STM Power, recently reached commercialization with a modular 55kW commercial unit. The manufacturer claims 30% electrical efficiency and 80% combined heat and power efficiency. A 25 kW Beta (field test unit) was installed at a WWTP in Corvallis (Oregon, US) although it was not operated with cogeneration. The technology has been tested to withstand H2S and siloxanes. The only treatment is some water removal such that it’s not entering the machine fully saturated (prevent liquid droplets). The dew point should be – 7°C (20°F) below the gas temperature. The range for the gas temperature is 43-60°C (110-140°F). Compression not likely required for digester gas; generally 2 psi is sufficient (Alexandrian, 2004).

Corvallis applied to the Energy Trust of Oregon for a grant of $49,424 to install the 55 kW unit in full operation at the treatment plant. The estimated annual generation was about 409,530 kWh per year. It was planned that all energy (12% of the plant’s electricity needs) be used onsite. It was also planned that the project be owned and operated by a third-party developer under a shared-savings arrangement with Corvallis. The responsibilities of the developer included operations, maintenance, management and insurance on the facility (Energy Trust of Oregon, 2004).

3.3.7 Summary of Energy Recovery Systems

Pertinent information regarding ERSs are summarized in Table 3-15. Boilers are still one of the most cost-effective methods for recovering energy, having no gas pretreatment requirements, high thermal recovery efficiency and low capital and operating costs. The drawback to boilers is that it recovers thermal energy only, and so may not be as useful in the warmer temperate or tropical climates. Fuel cells are the most capital intensive and have the highest annual O&M expenses. With time, the newer systems such as fuel cells and microturbines may decline in cost.
**Table 3-15. Summary of the Various Energy Recovery Systems.**

<table>
<thead>
<tr>
<th>ERS</th>
<th>Gas Pretreatment Requirements¹</th>
<th>Thermal Efficiency</th>
<th>Electrical Efficiency</th>
<th>Capital Cost²</th>
<th>O&amp;M Costs²</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler</td>
<td>None</td>
<td>75-80%</td>
<td>N/A</td>
<td>$100/kWh</td>
<td>$0.56/MJ</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>($23,500/hp)</td>
<td>($470/hp-yr)</td>
</tr>
<tr>
<td>Gas Engine Cogeneration</td>
<td>A,B,C</td>
<td>45-50%</td>
<td>30-35%</td>
<td>$1,000/kW-</td>
<td>$13/kW/yr</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$1,250/kW</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas Engine Direct Drive</td>
<td>A,B,C</td>
<td>~46%</td>
<td>26-28%</td>
<td>$22/m3/d</td>
<td>$1.10/m3/d</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>($82,036/MGD)</td>
<td>($4,219/MGD)</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Microturbine</td>
<td>A,C</td>
<td>35-40%</td>
<td>24-28%</td>
<td>$1450/kW</td>
<td>$130/kW/yr</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Phosphoric Acid Fuel Cells (PAFC)</td>
<td>A,B</td>
<td>~37%</td>
<td>~40%</td>
<td>$6250/kW</td>
<td>$250/kW/yr</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

¹ A = Water Vapor; B = H₂S; C = Siloxane
² Capital and operating & maintenance costs are reported in US Dollars (USD)

### 3.4 Life Cycle Analysis applied to Wastewater Treatment

LCA is an analytical tool used to determine the environmental and economic effects of a product, process or project at every stage of its existence. In an LCA, an inventory of the environmental emissions and resource use of an activity are determined and their subsequent environmental impacts are determined. This is a four-step process that is comprised of the following steps (Curran, 1996):

11. **Goal Definition and Scoping:** The initial step establishes the objective of the study and defines the boundaries of the scope and study limitations. The phase(s) of the life cycle are determined, and in this study will include the construction, operation, and demolition phases for WWTPs. Lastly in this step, the basis for the assessment is defined, and is referred to as the functional unit, such as a unit per volume of wastewater treated, (or perhaps as a unit per person equivalents).

12. **Inventory Analysis:** All data inputs and outputs for the system (digester and energy recovery processes) for the scope defined in Step 1 are collected and calculated so that similar systems may be analyzed and compared. The inputs and outputs generally include resource use and pollutant emissions. Information for this step can be through empirical data collected on-site, through established emission factors data or through simulation models.

13. **Impact Analysis:** This step in the LCA assigns some value to the effect which the process or project (anaerobic digestion plus energy recovery) will have on the environment. According to Curran (1996), the effects evaluated can include Global Warming Potential (GWP), Acidification Potential, Resource Depletion, Eutrophication, Human Toxicity, Eco-toxicity, and Photo-oxidant Creation Potential. Such effects are usually determined through weighting and normalizing the emissions found in Step 2.

14. **Improvement Analysis:** The final step in LCA is to evaluate the results of the impact analysis, and use of sensitivity analysis to determine the range of outcomes expected
over a specified range of inputs, followed by identification and evaluation of the opportunities to reduce environmental impacts.

The LCA method has several challenges to maintain objectivity in the analysis. The first challenge occurs when defining the scope. The scope of the LCA should address all substantial impacts of an economic activity throughout its life (from ‘cradle to cradle’). This becomes difficult when trying to assess indirect impacts of an activity such as energy and chemical production. Defining the scope should therefore be an iterative process complete with sensitivity analysis to ensure all significant impacts are addressed. A few hybrid approaches have been created that incorporate both empirical data and the EIO model. The empirical data is used to determine the impacts of the main components of the activity under analysis while an augmented EIO model is used to calculate pollution emissions and resource use generated by the indirect activities (MacLean et al., 2003; Hwang and Hanaki, 2000). Another challenge researchers often encounter with empirical data are gaps in the data or unreliable data. To alleviate these problems, emissions models have been created by STOWA a Dutch-based foundation for applied water research (the DEMAS or Designing and Evaluation Model for Wastewater treatment Scenario’s) and by Chalmers University (the Life Cycle Analysis Inventory Tool), which were used in two surveyed studies (Mels et al., 1999; Vidal et al., 2002).

Although, many procedures have been created to deal with challenges of the Life Cycle Assessment, LCA is still not a perfect analytical tool. Therefore, it is important to use LCA in collaboration with other decision-making tools to arrive at the best alternative for a situation.

Emmerson et al. (1995) performed the first LCA-based WWTP study in 1995. This study, unlike most LCA-based studies to date, focused mainly on the construction and demolition phases. It also analyzed the material and energy use in the operational phase of the WWTPs. Since then, there have only been a few LCA-based wastewater treatment studies that have included the construction phase (Lundin et al., 2000; Tillman et al., 1998). Only one other study surveyed included both the construction and demolition phase in the analysis (Hwang and Hanaki, 2000). There are two reasons given for not including the construction and demolition phases. Firstly, the emissions are insignificant when compared to the emissions from the use phase of the WWTP (Suh and Rousseaux, 2002). Secondly, even if these emissions were significant, they represent “impulse emissions” that are related to construction practices not the treatment type utilized in the WWTP (MacLean et al., 2003).

The LCA-based studies that addressed the operational phase focused on different aspects of the wastewater treatment process. Lundin et al. (2000) gave thorough literature reviews of wastewater treatment LCA-based studies up to the point of their studies. A few of those surveyed addressed biosolids treatment. Mels et al.(1999), for example, applied an LCA approach to different aerobic wastewater treatment configurations in the Netherlands and compared them based on chemical requirements, land use, and sludge production. Lundin et al. (2000) also addressed biosolids, however used a much larger scope to compare wastewater collection and treatment systems. In doing so, Lundin et al. (2000) showed the impact boundaries have on LCA results. Lundin, however, did not address the production or treatment of biogas, nor did any of the studies she surveyed in her literature review.
Many studies that have been published since Lundin et al. (2000) have addressed aspects of the wastewater treatment system that are beyond the scope of this research, such as the collection system, nitrogen removal, urine separation and primary sewer solids treatment (Vidal et al., 2002; Mulder, 2003; Maurer, 2003; Gouda et al., 2003). One study performed by Bridle and Skrypski-Mantele (2000) mentions the possibility of energy production from sludge through fuel creation or incineration, but does not analyze an actual ERS to determine the actual impacts.

Suh and Rousseaux (2002), Clauson-Kaas et al. (2001), Hwang and Hanaki (2000) and Bjorklund et al. (2001) all included biogas production in their analysis. In the study by Suh and Rousseaux (2002), five different sludge treatment techniques were assessed for a specific treatment facility in France. One of the scenarios analyzed incorporated anaerobic digestion. The study concluded that using anaerobic digestion was the least environmentally harmful scenario, but did not include energy recovery from biogas. The study done by Bjorklund et al.(2001), however, was only a Life Cycle Inventory (LCI) of one plant in Sweden. Similarly, the study by Hwang and Hanaki (2000) was an inventory of CO2 emissions of different wastewater treatment options in Japan. A full LCA was performed by Clauson-Kaas et al. (2001), however, it was for a specific WWTP in Denmark that only used co-generation gas engines for energy recovery.

These studies are good templates to start with when applying the LCA method to North American WWTPs. The results of most of the studies mentioned, however, are only applicable to the region in which the assessment was performed so cannot be used as a means of comparison (Suh and Rousseaux, 2002; Vidal, 2002; Clauson-Kaas et al., 2001). This is due to differences in electricity production types used, materials used, proximity to water and resources, and climate (Roelveld et al., 1997; Enviromega et al., 2003). These differences affect both the inventory analysis, as well as the weighting and normalization factors in the impact assessment. One study in Australia was able to adjust some of the weighting factors used in Europe to ones that would be more applicable to an Australian treatment plant (Beavis and Lundie, 2003). In their case, the application of biosolids to land has a much different impact than in Sweden where it will be illegal to land apply biosolids by 2005 (Pettersson, 2001).

Of the surveyed LCA-based studies, none applied to North American WWTPs. Instead, WWTP comparisons to date have examined a specific aspect of treatment systems, greenhouse gas emissions, energy use, and cost requirements (Enviromega Inc. et al., 2003). Barton and Atwater (2002), for example, looked specifically at nitrous oxide (N2O) in a WWTP with energy recovery from incineration.

Since Emmerson et al.’s first WWTP study in 1995, there have been many WWTP studies performed, however none surveyed has analyzed the environmental impacts of different ERSs from anaerobically digested biogas in North America. Thus the objective of this aspect of the project was to provide North American decision makers with a tool to help determine which, if any ERS is best for their facility on a life cycle basis.
3.5  U.S. and Canadian Regulations pertaining to Anaerobic Digestion and Energy Recovery

3.5.1  Air Emissions

3.5.1.1  Clean Air Act Amendments of 1990 (U.S.)
This is a broad set of regulations governing control of air pollution. Major elements of the 1990 CAAA included:

- Establishment primary and secondary National Ambient Air Quality Standards for six pollutants designated as “criteria” pollutants;

- Establishment of technology-based emission control standards called New Source Performance Standards, applicable to new modified or re-constructed stationary emission sources;

- Establishment of health-based National Emission Standards for Hazardous Air Pollutants (NESHAPs); and

- Emission regulations for areas with clean air for Prevention of Significant Deterioration.

The Amendments contain seven major titles dealing with specific issues, namely:

- Title I: Urban Smog including ground-level ozone from reactions of N$_2$O and VOCs

- Title II: Mobile sources, primarily carbon monoxide and ozone from hydrocarbon-fueled vehicles;

- Title III: Air Toxics are those compounds posing significant health risks, for which the act provides technology-based standards for major emitters;

- Title IV: Acid Rain, caused by sulfur and NOx emitted to the atmosphere, then returning to the earth’s surface as wet or dry deposition;

- Title V: Permits issued to sources of air pollution to ensure compliance with the regulations;

- Title VI: Ozone-depleting Chemicals, such as chlorofluorocarbons (CFCs) and hydrochlorofluorocarbons (HCFCs) which contribute to ozone “holes” in the upper atmosphere;

- Title VII: Enforcement issues allowing for imposition of civil and criminal penalties to promote compliance.

The six criteria pollutants include CO, lead (P), NOx, O$_3$, SO$_2$ and fine particulate matter defined as PM-10. Criteria pollutants which could affect ERSs operating on digester gas include, NOx, and SO$_2$ (if the biogas has particularly high levels of H$_2$S). In properly operating engines, CO emissions would not likely be of concern.

Title I may most affect wastewater treatment facilities’ choice of ERS in O$_3$ non-attainment areas. Combustion-based processes such as boilers, engine-generators and direct drive units may emit excessively high levels of NOx, which would not be favored in an O$_3$ non-attainment area.
Such regulations would favor the installation of low emission processes such as microturbines (which produce low NOx emissions compared to other combustion technologies) and fuel cells (which do not use combustion for energy production).

A limited number of hazardous air pollutants, including formaldehyde, acetaldehyde and acrolein, may also be emitted from combustion processes. The extent to which the contaminants are controlled is subject to air quality effects monitoring.

Potential web sites of interest include:

URL: [http://www.epa.gov/oar/caa/contents.html](http://www.epa.gov/oar/caa/contents.html)

URL: [http://www.epa.gov/oar/oaqps/greenbk/index.html](http://www.epa.gov/oar/oaqps/greenbk/index.html)

Title 16, Chapter 12- Federal Regulation and Development of Power

Subchapter II: Regulation of Electric Utility Companies Engaged in Interstate Commerce

Of particular interest is Sec 824a -3 – Cogeneration and Small Power Production, which regulates the selling of power to and from cogeneration and small power production (<80MW) facilities, which includes various alternate power production (e.g. wastes, biomass). This set of regulations would appear to impact WWTPs that are not using the thermal or electrical energy produced on-site, but are instead are selling the energy off-site.

URL: [http://www4.law.cornell.edu/uscode/16/ch12.html](http://www4.law.cornell.edu/uscode/16/ch12.html)

3.5.1.2 Kyoto Protocol (1997)

This is an agreement to curtail gaseous emissions contributing to global warming, also referred to as the greenhouse effect. The agreement requires a 5-8% reduction in national CO2-equivalent emissions from levels in 1990 by 2008-2012. This applies to six greenhouse gases (GHG): CO2, CH4, N2O, HFCs, perfluorocarbons (PFC), and sulfur hexafluoride (SF6). The Kyoto Protocol was ratified in Canada in 2002, with a 3.3% emission reduction requirement. At the time of this report preparation, the Kyoto Protocol was not ratified by the U.S. Congress. Potential web sites of interest include:

URL: [http://unfccc.int/resource/docs/convkp/kpeng.html](http://unfccc.int/resource/docs/convkp/kpeng.html)

URL: [http://unfccc.int/resource/convkp.html](http://unfccc.int/resource/convkp.html)

Aerobic biological wastewater treatment processes produce the GHG CO2, and may possibly emit the more potent gas N2O, depending on the capacity for nitrification and denitrification. CO2 emissions from municipal wastewater treatment are not now included in national inventories because the carbon source is considered short-term anthropogenic in nature (as opposed to the fossilized carbon of coal and petroleum). Anaerobic wastewater treatment processes produce both CH4 and CO2. All ERSSs involving combustion will produce the GHG CO2. The Kyoto Protocol could impact future regulations, particularly if trading of carbon credits is instated. For example, GHG credits could become a driving force for WWTPs to consider using more
digester gas on-site to minimize purchases of retail electricity. The reduction in purchased electricity translates into a GHG credit for reducing the quantity of fossil fuel required for power generation. Other industries may purchase the credits from the treatment facilities.

3.5.1.3 Ontario Regulation196/01 Airborne Contaminant Discharge Monitoring and Reporting (June 14, 2001). Province of Ontario, Canada

This regulation amends the first such legislation in Ontario promulgated in May 2001 as O.R. 127/01. The act stipulates a number of contaminants that must be monitored and reported on in the province, not just by industry, but also by large establishments such as shopping centers, commercial and educational buildings and hospitals. The Act defines three sectors, including power generation, Large Sources and Small Sources. It defines criteria for reporting information and frequency based on these sectors. Three hundred and fifty-eight contaminants in total have been identified for inclusion in the reporting schedules. The first category is called the Criteria Air Contaminants and GHG, and includes several contaminants listed as Criteria Air Pollutants in the U.S. CAAA of 1990. These include CO, SO2, NOx and total PM (PM_{10} and PM_{2.5}). The other contaminants are CO_{2}, HFC-134A, CH_{4}, N_{2}O, and VOCs. A second group of contaminants, which are listed due to toxic or hazardous properties, have designated release-based thresholds for reporting. The third group is an overlap with the contaminants listed in the federal National Pollutant Release Inventory (similar to compounds in the Toxics Release Inventory in the U.S.), and consists of 268 contaminants.

WWTPs can be considered in the categories of large and small sources, and so could be required to monitor and report on emission rates, in most cases on the Criteria Air contaminants and GHGs.

Ontario Regulation 397/01 Governing emissions trading under the Environmental Protection Act


This regulation and supporting trading code establishes the regulatory framework in Ontario for Emissions Trading by regulating NOx and SO2 emissions and setting out rules for trading of emission allowances and emission reduction credits. The regulation came into effect in December 31st, 2001 for Ontario Power Generation, but as of 2004 applies to Independent Power Producers as well. The web site of interest is:

URL: [http://www.ene.gov.on.ca/envision/air/etr/index.htm](http://www.ene.gov.on.ca/envision/air/etr/index.htm)

The regulations are available at:


The impact of NOx emission restriction on the project was discussed above. SO2 can form with the combustion of sulfur compounds in the biogas, particularly H2S. H2S is also a gas compound targeted for removal during gas pretreatment for the protection of the gas handling equipment. Therefore, this can have an impact on the choices for GPSs, and by improving the quality of the treated gas it may also increase the choices for ERSs.
3.5.2 Regulations Impacting Biosolids Management

3.5.2.1 US EPA Title 40 Code of Federal Regulations Part 503 (40 CFR 503)
This set of regulations establishes the minimum national standards for the use and disposal of sewage and non-sewage sludge in the U.S. Included are limits for metal and trace substance concentrations, pathogen reduction and vector attraction. Limits on application rates and minimum distance from designated sites are also included.

URL: http://www.epa.gov/owm/mtb/biosolids/503pe/index.htm

Of particular interest to this project were the regulations for achieving Class A level for biosolids use. The most common type of anaerobic digestion used at WWTPs is mesophilic, which can achieve at most, Class B level, and it is limited to a small number of applications. To achieve the designation of Class A biosolids, advanced digestion methods such as thermophilic digestion, TPAD or a combination thereof, must be used. Therefore, many WWTPs may be considering an upgrade to their digestion process in order to increase flexibility in biosolids application options. Alternatively, municipalities may consider moving from anaerobic digestion as a stabilization process to other processes such as alkaline stabilization.

Individual states may also establish their own regulations on biosolids use and disposal in addition to the minimum federal requirements. The following link leads to regulations for 28 states:

http://biosolids.policy.net/states/statereg.vtml?PROACTIVE_ID=cecfedc8e6e8cacec7c5eef3e5e55eebce
ce7e67e8e56e

3.5.2.2 Nutrient Management Act, 2002 – Ontario Regulation 267/03
This set of regulations governs land application standards of sewage and non-sewage biosolids in the Province of Ontario, Canada. Proximity to designated areas such as wells, crops, and surface water are regulated. Application rates, analysis and quality standards such as metal and E. coli concentration limits are stated. This regulation impacts WWTPs in Ontario in a way similar to 40 CFR 503 for the U.S. as described above. The regulations do not differentiate between Class A and Class B quality biosolids. The link below leads the detail information regarding the Nutrient Management Act, 2002.

URL: http://www.e-laws.gov.on.ca/DBLaws/Source/Regs/English/2003/R03267_e.htm

3.5.2.3 Guidelines for the Utilization of Biosolids and other Wastes on Agricultural Land (1996) (Supplement to Ontario Regulation 347 under the Environmental Protection Act)
These guidelines for Ontario precede the Nutrient Management Act of 2002. The focus is on agricultural application of biosolids, with a distinction between anaerobically digested sewage sludge, and other sewage sludge. Though these are guidelines and not regulations, they may still have an impact on biosolids use and disposal, as they include guidelines for pH, and concentrations of elements such as boron and sodium, that are not covered in Nutrient Management Act.
In Canada, biosolids use and disposal are under provincial jurisdiction. Regulations for the Province of Ontario have been discussed above. Other provinces have their own set of guidelines or regulations set out by their respective ministry of environment (they may or may not be enforced by law). Province of Quebec, for example, has criteria presented in the following link (Section 3.5.2.4).

3.5.2.4 Interim Criteria for the Reclamation of Fertilizing Residuals (November 2002)
SURVEY RESULTS

Mesophilic anaerobic processes are the predominant form of anaerobic wastewater solids digestion. Digestion processes listed in Table 4-1 other than single- and two-stage mesophilic digestion are considered as innovative anaerobic digestion processes. While the detailed survey forms were sent to other facilities, particularly the ones with processes deemed innovative, responses were not forthcoming in the available time period.

The distribution of responding treatment plants with different types of ERSs is provided in Table 4-2. Boilers were the most common form of energy recovery from biogas, followed by engine-generators. These systems represent the more traditional digester gas ERSs. Microturbines and direct drive units appear to have growing acceptance as ERSs.

<table>
<thead>
<tr>
<th>Table 4-1. Types of Digestion Systems in Survey Responses.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Type of Anaerobic Digestion System</td>
</tr>
<tr>
<td>Mesophilic – High Rate Single Stage</td>
</tr>
<tr>
<td>Mesophilic – High Rate Two Stage (mixed/unmixed secondary digesters)</td>
</tr>
<tr>
<td>Thermophilic – High Rate Single Stage</td>
</tr>
<tr>
<td>Thermophilic – High Rate Multi Stage</td>
</tr>
<tr>
<td>Temperature phased (M/T or T/M)</td>
</tr>
<tr>
<td>Acid-Gas phased</td>
</tr>
<tr>
<td>TOTAL</td>
</tr>
</tbody>
</table>

^a data requested but no surveys returned
M = Mesophilic, T = Thermophilic

<table>
<thead>
<tr>
<th>Table 4-2. Types of Energy Recovery Systems in Survey.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Type of Energy Recovery System</td>
</tr>
<tr>
<td>Boilers</td>
</tr>
<tr>
<td>Engine-generators</td>
</tr>
<tr>
<td>Microturbines</td>
</tr>
<tr>
<td>Fuel Cells (Molten carbonate or phosphoric acid)</td>
</tr>
<tr>
<td>Direct Drive</td>
</tr>
<tr>
<td>Other</td>
</tr>
<tr>
<td>TOTAL</td>
</tr>
</tbody>
</table>

^a 2 plants surveyed use biogas in fuel cells. Additionally, Terminal Island uses natural gas only in fuel cells, King County in Seattle WA has fuel cells (plant was visited by did not complete a survey)
^b Includes 1 flare only; 1 sending biogas through pipeline; 1 steam turbine; 1 gas turbine, 1 combined cycle turbine and 1 pelletizing burner
^c Total number of systems is greater than the number of plants surveyed because some plants have more than one type of energy recovery system.

Thermal and electrical efficiencies for the different types of ERSs determined in the survey responses, relative to literature values, are summarized in Table 4-3. In general, the values determined in this study agree well with values reported in the technical literature.
Table 4-3. Comparison of Energy Recovery Efficiencies in Survey Responses and Literature Values.

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Boilers</td>
<td>75-80%</td>
<td>80%</td>
<td>---</td>
<td>---</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td>Gas Engines</td>
<td>45-50%</td>
<td>40.5%</td>
<td>30-35%</td>
<td>35.8%</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td>Direct Drives</td>
<td>~46%</td>
<td>33%</td>
<td>---</td>
<td>---</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td>Microturbines</td>
<td>35-40%</td>
<td>70%</td>
<td>24-28%</td>
<td>27.5%</td>
<td>26-28%</td>
<td>33%</td>
</tr>
<tr>
<td>Fuel Cells (PAFC)</td>
<td>~37%</td>
<td>N/A</td>
<td>~40%</td>
<td>29%</td>
<td>---</td>
<td>---</td>
</tr>
</tbody>
</table>
CHAPTER 5.0

DEVELOPMENT OF LCAMER NON-MONETARY AND COSTING FUNCTIONS FOR LIFE CYCLE ASSESSMENT FRAMEWORK

5.1 Introduction to Function Development

This section outlines the general framework for calculating the life cycle cost of energy recovery from anaerobically treated municipal sludge. Before generalized cost functions for the framework can be developed, non-monetary functions (technical functions) pertaining to the involved processes must first be developed. The processes associated with energy recovery from anaerobic sludge digestion are presented and the variables and functions involved with operation are presented for each process. The cost functions are then presented, from the most general function (life cycle cost) to more specific cost functions.

5.2 Technical Functions for the LCAMER Spreadsheet Model

The non-monetary (technical) functions have four variable types: input, operational, intermediate, and output. Input variables are those that are fully independent of the given unit process (e.g. anaerobic digestion). Operational variables are those that may be independently controlled. Though similar to input variables in that their values are not dependent on those of other variables in the same unit process, operational variables differ from input variables in that the former are not independent of the unit process; they are very much an integral part of that process. Though not operational variables are not variables per se, as they pertain to the application for a given facility, they are variables in that they are site-specific and therefore vary from one facility to the next. Intermediate variables are dependent to input and/or operational variables, and independent to output variables. Output variables are dependent upon the input, operational, and/or intermediate variables within the same unit process. Note that output variables for one process may be input variables for another process further downstream.

5.2.1 Anaerobic Digestion Process

Table 5-1 is an example of the non-monetary variables involved with the anaerobic digestion process.
Table 5-1. Non-Monetary Anaerobic Digestion Variables.

<table>
<thead>
<tr>
<th>Input Variables</th>
<th>Operational Variables</th>
<th>Intermediate Variables</th>
<th>Output Variables</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sludge flow rate, scf/d (m^3/d)</td>
<td>Anaerobic digestion process</td>
<td>% VS Reduction</td>
<td>Sludge flow rate_{out}, scf/d (m^3/d)</td>
</tr>
<tr>
<td>Sludge type, (PS/WAS)</td>
<td>Digester shape</td>
<td>SRT, d</td>
<td>TS_{out}, lb/gal (g/m^3)</td>
</tr>
<tr>
<td>TS_{in}, lb/gal (g/m^3)</td>
<td>total digester volume, scf (m^3)</td>
<td>HRT, d</td>
<td>VS_{out}, lb/gal (g/m^3)</td>
</tr>
<tr>
<td>VS_{in}, lb/gal (g/m^3)</td>
<td>Digester sludge temp, °F (°C)</td>
<td>scf biogas/lbVS (m^3 biogas/kgVS)</td>
<td>Biogas production rate scf/d (m^3/d)</td>
</tr>
<tr>
<td></td>
<td>dry pH buffer dosage, lb/gal (g/m^3)</td>
<td>dry pH buffer addition, lb/d (kg/d)</td>
<td>biogas composition</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % methane^1</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % CO2</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % H_2S</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % H_2</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % N_2</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Biogas % water vapour</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>biogas % other</td>
</tr>
</tbody>
</table>

^1 Biogas composition for all compounds is % by volume (v/v).

5.2.1.1 Volatile Solids Reduction

Estimation of the daily biogas production rate is dependent on the extent of the VS Reduction of the wastewater solids delivered to the anaerobic digester. The VS Reduction is dependent on a number of parameters including the SRT in the digester, the proportions of primary sludge and waste activated sludge (WAS), and operating temperature of the digester.

An initial semi-empirical steady state model was derived to predict the VS Reduction. The derivation was an important step forward in this study, in terms of simplicity for the application in this spreadsheet model, as well as in very good accuracy (compared to results with the IWA ADM1 model; Batstone et al., 2002). As initially developed, the model worked successfully with staged thermophilic and mesophilic processes. The model had a limitation, however, in that it did not successfully allow for the formation of soluble products in the first of two or more stages, such as an acid-gas system. As a result, the model was re-developed to account for formation of the soluble products (Kalogo et al., 2005). The result appears in Equation 5-1. The values of the parameters for calibrating the VSR model are presented in Appendix D.

\[ \text{VS}_i = \sum \left[ \frac{\text{VSS}_{i,j}}{1 + \text{SRT} \times K_{1,j} + \text{VS}_{i,j}} \times \left( 1 + \text{SRT} \times K_{1,j} \right) \frac{\text{S}_{i,j}}{1 + \text{SRT} \times K_{1,j}} \right] \]

(5-1)

Where:

\( F_{v,j} \) = Conversion factor from volatile suspended solids (VSS) to chemical oxygen demand (COD) units, lb\textsubscript{COD}/lbVSS (kg\textsubscript{COD}/kgVSS)

\( K_{1,T,c,j} \) = First order substrate consumption for sludge \( j \) at temperature \( T \) and process condition \( c \), d\(^{-1}\)

\( K_{h,T,c,j} \) = First order hydrolysis constant for sludge \( j \) at temperature \( T \) and process condition \( c \), d\(^{-1}\)
\( j \) = Sludge type: \( j = 1 \) for primary sludge; \( j = 2 \) for WAS

\( S_{H,i-1,j} \) = Soluble concentration of the constituents entering stage \( i \), lb\( _{\text{COD/gal}} \) (kg\( _{\text{COD/m}^3} \))

\( \text{SRT}_i \) = Design sludge retention time of stage \( i \), d

\( \text{SRT}^*_i \) = Hypothetical sludge retention time of stage \( i \), d

\( \text{VS}_i \) = Concentration of VS leaving stage \( i \), lb\( _{\text{VS/gal}} \) (kg\( _{\text{VS/m}^3} \))

\( \text{VSS}_{i-1,j} \) = VSS concentration of sludge \( j \) entering stage \( i \), lb/gal (kg/m\(^3\))

Then the VS Reduction is given by:

\[
\text{VSR}(\%) = 100 \left[ \frac{\text{VS}_0 - \text{VS}_i}{\text{VS}_0} \right]
\]  \hspace{1cm} (5-2)

Where:

\( \text{VSR} \) = Volatile solids reduction, %

\( \text{VS}_0 \) = Concentration of VS entering stage 1, lb\( _{\text{VS/gal}} \) (kg\( _{\text{VS/m}^3} \))

\( \text{VS}_i \) = Concentration of VS leaving stage \( i \), lb\( _{\text{VS/gal}} \) (kg\( _{\text{VS/m}^3} \))

### 5.2.1.2 Biogas Production Yield

The biogas production yield is based on the VS Reduction of the raw solids multiplied by the biogas yield factor of VS added.

\[ \text{GPR} = Y_{\text{biogas}} \times \text{VSR} \]  \hspace{1cm} (5-3)

Where:

\( \text{GPR} \) = Gas production rate, scf/lb\( _{\text{VS added}} \) (m\(^3\)/kg\( _{\text{VS added}} \))

\( \text{VSR} \) = Volatile solid reduction, %

\( Y_{\text{biogas}} \) = Biogas yield, 14.42 scf/lb (0.9 m\(^3\)/kg)

### 5.2.1.3 Daily Flow of Wastewater

The default estimation of the wastewater flow rate is based on the population (number of inhabitants) served by the wastewater treatment system, multiplied by factors determined from the returned survey questionnaires for this project. Flow rates were determined for separate (without stormwater) and combined (with stormwater) sewer systems.

\[ Q_W = Z_{\text{flow}} \times \text{AFPP} \times P \]  \hspace{1cm} (5-4)

Where:

\( Q_W \) = Daily wastewater flow, MGD (m\(^3\)/d)

\( P \) = Population served, number of people
AFPP  = Average flow per person, gpd/cap (m³/cap-d), based on the survey results as follows:
    AFPP_{Separated} = 88.37 gal/cap-d (0.3345 m³/cap-d, R² = 0.9526)
    AFPP_{Combined} = 130.40 gal/cap-d (0.4936 m³/cap-d, R² = 0.9099)

Z_{flow}  = Conversion factor for flow, 10⁻⁶ MGD/gpd (1 m³/d / 1 m³/d)

5.2.1.4 Primary Sludge Flow
The volumetric flow rate of primary sludge is estimated using the population served, and factors adjusting the volume based on the operation of primary settling tanks. A per capita primary sludge production rate of 2.51x10⁻⁵ unit volume grit per unit volume raw wastewater was derived from Satin and Selmi (1999). The model user must enter the volumetric flow rate of chemical coagulants used in the primary clarification step.

\[
Q_{gp} = \alpha \times [Q_{W} \times Z_{flow} - (Q_{W} \times R_{G/W} \times Z_{flow}) + V_{chemical}]
\]  \hspace{1cm} (5-5)

Where:
- \(Q_{SP}\) = Primary sludge flow, gpd (m³/d)
- \(Q_{W}\) = Daily raw wastewater flow, MGD (m³/d)
- \(R_{G/W}\) = Volume ratio of grit to raw wastewater, 2.51x10⁻⁵ gal_gal_w (2.51x10⁻⁵ m³_gal_w/m³_w)
- \(V_{chemical}\) = Volume of chemical added per day, gpd (m³/d)
- \(\alpha\) = Proportion percentage coefficient, using the following values:
  - \(\alpha = 0\%\) (if no primary settler)
  - \(\alpha = 1\% \& V_{chemical} = 0\) (if natural settling)
  - \(\alpha = 3\%\) (if chemical enhancing primary treatment)
- \(Z_{flow}\) = Conversion factor for volume, 10⁻⁶ gpd/MGD (1 m³/d / m³/d)

5.2.1.5 Daily Tonne of Thickened Primary Sludge
The mass rate of thickened primary sludge produced is calculated from the volumetric flow rate of primary sludge, the density of the primary sludge, and the concentration of VS in the primary sludge.

\[
D_{TPS} = D_{SP} \times Q_{SP} \times C_{SP} \times Z_{weight}
\]  \hspace{1cm} (5-6)

Where:
- \(D_{TPS}\) = Daily tonne VS of primary sludge, ton/d (tonne/d)
- \(D_{SP}\) = Density of primary sludge, 8.345lb/gal (1,000kg/m³)
- \(Q_{SP}\) = Primary sludge flow, gpd (m³/d)
- \(C_{SP}\) = Concentration of thickened primary sludge entering the digester, %VS
- \(Z_{weight}\) = Conversion factor for weight, 2000 lb/ton (1000kg/tonne)
5.2.1.6 Biological Sludge Flow

In the spreadsheet model user must enter five parameters for this calculation, and may instead choose to directly enter the thickened sludge flow and the percent VS in the thickened sludge (%VS\textsubscript{th}), and not use the results of this calculation.

The default thickened biological sludge flow is calculated using the following expressions. The daily mass of biological VSS wasted is estimated as:

\[
VSS\text{\textsubscript{wasted}} = \frac{\text{MLSS} \cdot V_{\text{aeration}} \cdot \%\text{VSS} \cdot Z_{\text{volume}}}{\text{SRT}_{\text{aerobic}} \cdot Z_{\text{weight}}} \tag{5-7}
\]

Where:
- \(VSS\text{\textsubscript{wasted}}\) = Daily mass of biological VSS wasted, lb/d (kg/d)
- \(\text{MLSS}\) = Concentration of mixed liquor suspended solids, mg/L
- \(V_{\text{aeration}}\) = Aeration basin volume, gal (m\(^3\))
- \(\%\text{VSS}\) = Percent Fraction of Thickened Biological Sludge
- \(\text{SRT}_{\text{aerobic}}\) = SRT of the biological treatment system, d
- \(Z_{\text{weight}}\) = Conversion factor for weight, 454,000 mg/lb (10\(^6\) mg/kg)
- \(Z_{\text{volume}}\) = Conversion factor for volume, 3.785L/gal (1000L/m\(^3\))

The percent VS in the thickened sludge (%\(\text{VS}_{\text{th}}\)) is estimated from:

\[
\%\text{VS}_{\text{th}} = \%\text{VSS} \cdot \%\text{TS}_{\text{th}} \tag{5-8}
\]

Based on:
- \(\%\text{TS}_{\text{th}}\) = Total solids concentration in the thickened sludge (entered by the user).

The thickened sludge flow (\(Q_{\text{wsb}}\)) is then determined from:

\[
Q_{\text{wsb}} = \frac{VSS\text{\textsubscript{wasted}}}{\%\text{VS}_{\text{th}} \cdot Z_{\text{concen}}} \tag{5-9}
\]

Where:
- \(Q_{\text{wsb}}\) = Thickened sludge flow, gal/d (m\(^3\)/d)
- \(VSS\text{\textsubscript{wasted}}\) = Daily mass of biological VSS wasted, lb/d (kg/d)
- \(\%\text{VS}_{\text{th}}\) = Percentage VS in thickened sludge, %\(\text{VS}_{\text{th}}\)
- \(Z_{\text{concen}}\) = Concentration of VS in the thickened sludge flow, 0.0835lb/gal/%VS (10kg/m\(^3\)/%VS)

These estimates assume that the mass of VS leaving with the effluent stream is insignificant compared to the VS in the thickened sludge stream.
5.2.1.7 Daily Tonne of Biological Sludge

The mass flow rate of biological solids wasted to the digester daily is calculated as a product of the biological solids volumetric flow rate and the concentration of VS in the wasted flow.

\[
DTBS = DSB \times QSB \times C_{SB} \times Z_{volume} / Z_{weight} \quad (5-10)
\]

Where:
- \( DTBS \) = Daily tonne VS of biological sludge, ton/d (tonne/d)
- \( DSB \) = Density of biological sludge, 1,000,000mg/L (1000kg/m\(^3\))
- \( QSB \) = Biological sludge flow, gpd (m\(^3\)/d)
- \( C_{SB} \) = Concentration of biological sludge entering the digester, %VS
- \( Z_{weight} \) = Conversion factor for weight, 1.102x10\(^6\) mg/ton (1000kg/tonne)
- \( Z_{volume} \) = Conversion factor for volume, 3.785L/gal (1m\(^3\)/m\(^3\))

5.2.1.8 Daily Sludge Flow to Digesters

The total daily volumetric flow rate of solids to the digester is the sum of the primary and waste secondary sludge daily flow rates.

\[
QS = QSP + QSB \quad (5-11)
\]

Where:
- \( QS \) = Daily sludge flow to digesters, gpd (m\(^3\)/d)
- \( QSP \) = Primary sludge flow, gpd (m\(^3\)/d)
- \( QSB \) = Biological sludge flow, gpd (m\(^3\)/d)

5.2.1.9 Daily Tonne versus Treated

The daily mass of VS sent to the digester for treatment is the sum of the daily mass rates of volatile solids in the primary and secondary sludge streams.

\[
DT_S = DTP_S + DTBS \quad (5-12)
\]

Where:
- \( DT_S \) = Daily tonne VS treated, ton/d (tonne/d)
- \( DTP_S \) = Daily tonne VS of primary sludge, ton/d (tonne/d)
- \( DTBS \) = Daily tonne VS of biological, ton/d (tonne/d)

5.2.1.10 Working Volume of Digester

The digester working volume is calculated as the product of the digester SRT and the daily volumetric sludge flow rate to the digester.
V_{W} = SRT \times Q_{S} \quad (5-13)

Where:

V_{W} = \text{Working volume of digesters, gal (m}^{3}\text{)}
SRT = \text{Total sludge retention time of digesters, d}
Q_{S} = \text{Daily sludge flow to digesters, gpd (m}^{3}/\text{d)}

5.2.1.11 \hspace{1cm} \textbf{Total Volume of Digesters}

The total digester volume, used for costing purposes, is estimated from the digester working volume by adjusting the working volume with a scale-up factor to the total volume. The default value is 0.9 for working-to-total volume.

V_{T} = \frac{V_{W}}{R_{W/T}} \quad (5-14)

Where:

V_{T} = \text{Total volume of digesters, gal (m}^{3}\text{)}
V_{W} = \text{Working volume of digesters, gal (m}^{3}\text{)}
R_{W/T} = \text{Ratio working volume total volume, gal/gal (m}^{3}/\text{m}^{3}\text{)}.

A default value \text{ } R_{W/T} = 0.9 \text{ is assumed.}

5.2.1.12 \hspace{1cm} \textbf{Organic Loading Rate}

The digester organic loading rate is calculated as the daily mass rate of VS entering the digester divided by the working volume of the digester.

\text{OLR} = \frac{D_{TS}}{V_{W}} \quad (5-15)

Where:

\text{OLR} = \text{Organic loading rate, ton VS/gal-d (tonne VS/m}^{3}\text{-d)}
D_{TS} = \text{Daily tonne VS treated, ton VS/d (tonne VS/d)}
V_{W} = \text{Working volume of digesters, gal (m}^{3}\text{)}

5.2.1.13 \hspace{1cm} \textbf{Daily Raw Biogas Produced}

The daily volumetric flow rate of biogas produced by the digestion system is calculated as the product of the biogas yield and the mass of volatile solids sent daily to the digester.

Q_{RB} = GPR \times D_{TS} \times Z_{weight} \quad (5-16)

Where:

Q_{RB} = \text{Daily raw biogas produced, scf/d (m}^{3}/\text{d)}
GPR = \text{Gas production yield, scf/lb VS added (m}^{3}/\text{kg VS added)}
D_{TS} = \text{Daily tonne VS treated, ton/d (tonne/d)}

Cost-effective Energy Recovery from Anaerobically Digested Wastewater Solids
\[ Z_{\text{weight}} = \text{Conversion factor for weight, 2000lb/ton (1000kg/tonne)} \]

### 5.2.1.14 Heat Loss Calculations

#### Energy Needed for Digester Heating

The energy needed as heat for the digestion process includes both the heat required to raise the raw sludge to the digester operating temperature, as well as the heat required to make up for heat losses through the digester skin.

#### Energy Needed for Digester Feed Sludge Heating

\[ \begin{align*}
E_{\text{DH}} &= \frac{\rho_{\text{SC}} \times Q_S \times c \times [T_2 - T_1]}{Z_e \times \text{HEE}} \\
T_1 &= \text{Temperature of feed sludge to be treated °F (°C), value calculated by state and province using empirical relationship}
\end{align*} \]

Where:
- \( E_{\text{DH}} \) = Energy needed to raise the raw sludge to digester temperature Btu/d (MJ/d)
- \( \text{HEE} \) = Heat exchanger efficiency (%)
  - A default value of \( \text{HEE} = 80\% \) is assumed
- \( Q_S \) = Daily sludge flow to digesters gpd (m³/d)
- \( \rho_{\text{SC}} \) = Density of combined sludge lb/gal (kg/m³)
  - Assumed values are \( \rho_{\text{S primary sludge}} = 8.51 \text{ lb/gal (1020 kg/m³)} \) and \( \rho_{\text{S biological sludge}} = 8.39 \text{ lb/gal (1005 kg/m³)} \)
- \( c \) = Average specific heat of sludge Btu/lb/°F (J/kg/°C)
  - An assumed value of \( c = 1 \text{ Btu/lb/°F (4200 J/kg/°C)} \) is used.
- \( T_2 \) = Temperature of sludge from digester °F (°C) or digester 1 temperature °F (°C)
  - Assumed values of \( T_{2,\text{Meso}} = 95^\circ\text{F (35 °C)} \) and \( T_{2,\text{Thermo}} = 131^\circ\text{C (55 °C)} \) are used.
- \( Z_e \) = Conversion factor for energy, 1 Btu/Btu (10^6 J/MJ)

#### Energy Needed for Digester Surface Heat Loss

\[ \begin{align*}
E_L &= \frac{U \times A_C \times [T_2 - T_3] \times Z_t}{Z_e \times \text{HEE}} \\
T_3 &= \text{Temperature of sludge from digester °F (°C) or digester 1 temperature °F (°C)}
\end{align*} \]

Where:
- \( E_L \) = Energy loss Btu/d (MJ/d)
- \( \text{HEE} \) = Heat exchanger efficiency (%)
  - A default value of \( \text{HEE} = 80\% \) is assumed
- \( U \) = Heat transfer coefficient Btu/ft²-°F-h (W/m² °C)
  - An assumed value of \( U = 0.1410 \text{ Btu/ft²-°F-h (0.8 W/m².C)} \) is used.
AC = Total surface (including top and bottom) of digester ft² (m²)

T₂ = Temperature of sludge from digester (°C) or digester 1 temperature (°C)

Assumed values of \( T_{2\text{Meso}} = 95^\circ F \) (35 °C) and \( T_{2\text{Thermo}} = 131^\circ F \) (55 °C) are used.

T₃ = Average ambient temperature °F (°C) per jurisdiction

[Mean values by state and province are provided].

Zₜ = Conversion factor for time, 24h/d (86,400 J/d.W or s/d)

Zₑ = Conversion factor for energy, 1 Btu/Btu (10⁶ J/MJ)

**Temperature Adjustments for sludge temperatures and ambient temperatures**

Energy is required to raise the temperature of the sludge pumped to an anaerobic digester to the digester’s operating temperature. Energy may also be required to accommodate potential heat losses from the digester tank surfaces, especially in more northern States and Canada.

Two procedures were investigated to estimate the sludge input temperature if values are not readily available from model users. The first procedure is based on an empirical relationship, between ambient temperatures for the American states and Canadian provinces, and the temperature of sludge pumped to the digester. This empirical relationship is a linear function developed by using literature data. The empirical relationship derived in this case is:

\[
\text{Sludge temperature (°F or °C)} = x_1 \text{ Ambient temperature (°F or °C)} + x_2
\]

\[
x_1 = \text{Variable 1, 0.6071 for temperature in °F (0.607 for temperature in °C)}
\]

\[
x_2 = \text{Variable 2, 30.931 for temperature in °F (10.2 for temperature in °C)}
\]

Predicted sludge temperatures pumped to an anaerobic digester based on the linear empirical relationship are also shown in Appendix E for the US and Canada, respectively.

The ambient air temperature data can also be used to calculate digester tank surface heat losses. Depending on the material of construction and insulation factor, standard heat transfer equations can be used to estimate the heat loss based on thermal transfer resistance properties of the material of construction and the digester operating and ambient air temperatures.

The second procedure for calculating digester heating requirements is a more mathematical approach developed by Talati and Stenstrom (1990) for heat losses from wastewater aeration basins. This approach involves calculating:

15. 1. Heat Gains from a) Solar Radiation, b) Power Input, and c) Biological Reactions

16. 2. Heat Losses from d) Atmospheric Radiation, e) Surface Aeration, f) Surface Convection, g) Subsurface Aeration, and h) Tank Wall Convection/Conduction

With anaerobic digesters, the appropriate terms can be omitted, such as subsurface aeration. The limitations to the detailed mathematical approach are that it requires many site-specific factors as inputs, greatly increasing the difficulty and complexity for end-users. Project Team believed the
empirical approach was more appropriate for the intended users of the spreadsheet model, who may have limited information available to use the Talati and Stenstrom procedure.

### 5.2.1.15 Energy Needed for Digester Mixing

The electrical energy needed for keeping the contents of the digester mixed is calculated as the energy needed per volume of digestion tank (e_{DM}) multiplied by the digester working volume.

\[
E_{DM} = e_{DM} \times V_W \hspace{1cm} (5-19)
\]

Where:

- \(E_{DM}\) = Energy needed for digester mixing, \(W\)
- \(e_{DM}\) = Energy needed for digester mixing per volume of digester, \(0.0246\,W/\text{gal} (6.5\,W/m^3)\)
- \(V_W\) = Working volume of digesters, \(\text{gal} (m^3)\)

### 5.2.2 Digester Gas Pretreatment

A general schematic presentation of the gas pretreatment unit is presented in Figure 5.1. The technical variables for gas pretreatment are presented in Table 5.4. The incoming digester gas flow rate is an operational variable because the facility’s management must decide the amount of digester gas that is going to be used, for example, for building space heating.

<table>
<thead>
<tr>
<th>Input Variables</th>
<th>Intermediate Variables</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas process technology</td>
<td>% H\textsubscript{2}S removal</td>
</tr>
<tr>
<td>Chemical regeneration process</td>
<td>% CO\textsubscript{2} removal</td>
</tr>
<tr>
<td>Digester gas contaminant composition\textsubscript{1}</td>
<td>% Water Vapor removal</td>
</tr>
<tr>
<td>% Methane\textsubscript{1}</td>
<td>% Siloxanes removal</td>
</tr>
<tr>
<td></td>
<td>Chemical addition, dry lb/d (dry kg/d)</td>
</tr>
<tr>
<td></td>
<td>Used chemicals, dry lb/d (dry kg/d)</td>
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<tr>
<td></td>
<td>Regenerated chemicals, dry lb/d (dry kg/d)</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Operational Variables</th>
<th>Output Variables</th>
</tr>
</thead>
<tbody>
<tr>
<td>Digester gas flow rate\textsubscript{1}: scf/d (m\textsuperscript{3}/d)</td>
<td>Digester gas flow rate\textsubscript{2}</td>
</tr>
<tr>
<td></td>
<td>Digester gas composition\textsubscript{2}, Methane\textsubscript{2}</td>
</tr>
<tr>
<td></td>
<td>Chemical wasting rate in digester gas pretreatment system, dry lb/d (dry kg/d)</td>
</tr>
<tr>
<td></td>
<td>Chemical wasting rate in chemical regeneration system, dry lb/d (dry kg/d)</td>
</tr>
</tbody>
</table>
5.2.3 Biogas Technical Functions

5.2.3.1 Daily Raw Biogas for Energy Production

The energy in the raw biogas is calculated as thermal energy from boilers and as electrical energy from other processes as calculated below.

For thermal energy production

\[
Q_{RBTE} = \frac{Z_{time} \cdot C_{ERS} \cdot %CH_{4PB} \cdot ECE_{Ther} \cdot ECH_{CH} \cdot Z_{energy}}{\text{Daily raw biogas for thermal energy production, scf/d (m}^3/\text{d)}}
\]  \hspace{1cm} (5-20)

Where:

\(Q_{RBTE}\) = Daily raw biogas for thermal energy production, scf/d (m\(^3\)/d)
\[ Q_{RBEE} = \frac{Z_{time} \times C_{ERS}}{\%CH_{4RB} \times ECE \times EC_{CH4} \times Z_{energy}} \]  

Where:

- \( Q_{RBEE} \) = Daily raw biogas for electrical energy production with Stirling engines or steam turbines, scf/d (m³/d)
- \( Z_{time} \) = Conversion factor for time, 1d/d (86,400s/d)
- \( C_{ERS} \) = Capacity of energy recovery of Stirling engines or steam turbines, kW
- \( \%CH_{4RB} \) = Percentage of CH₄ in raw biogas, %  
  \textit{Assumed value is \%CH_{4RB} = 60\%}.
- \( ECE \) = Energy recovery efficiency, %
- \( EC_{CH4} \) = Energy content of methane, Btu/scf (MJ/m³)  
  \textit{Assumed value is EC_{CH4} = 961.5Btu/scf (35.8 MJ/m³)}.
- \( Z_{energy} \) = Conversion factor for energy, 1Btu/h/Btu/h (1000kJ/MJ)

For electrical energy production with Stirling engines or steam turbines

\[ Q_{DRBEE} = Q_{RBTE} + Q_{RBEE} \]  

Where:

- \( Q_{DRBEE} \) = Total daily raw biogas for energy production, scf/d (m³/d)
- \( Q_{RBTE} \) = Daily raw biogas for thermal energy production, scf/d (m³/d)
- \( Q_{RBEE} \) = Daily raw biogas for electrical energy production with Stirling engine or steam turbine, scf/d (m³/d)

5.2.3.2 Biogas Quality

When the quantity of biogas needing purification is determined, steps can be taken to select the purification steps. This effectively alters the biogas flow rate only if CO₂ removal is specified.
For H₂S and siloxane removals only, the purified biogas flow rate is virtually identical with the raw biogas flow rate entering the purification stages. Calculations are made to determine final CO₂, H₂S and siloxane concentrations in the purified biogas based on the user input concentrations for the raw biogas, and specified removal efficiencies. For example, if a facility were to send biogas offsite, after upgrading for CO₂ removal, the flow of purified biogas would be reduced by approximately 40 %, but the CH₄ content by volume would increase, say from 60 % to a higher quality of approximately 91 %. Biogas purified and sent offsite affects the operating costs for CO₂ removal in the economic functions.

5.2.3.3 Content of Purified Biogas

% CO₂ in purified biogas

\[
\%\text{CO}_2^{PB} = \%\text{CO}_2^{RB} \times [1 - \text{RCO}_2] 
\] (5-23)

Where:

\%\text{CO}_2^{PB} = \text{Percentage of CO}_2 \text{ in purified biogas, } \%
\%\text{CO}_2^{RB} = \text{Percentage of CO}_2 \text{ in raw biogas, } \%
\text{RCO}_2 = \text{CO}_2 \text{ Pretreatment Efficiency, } \%

ppm H₂S in purified biogas

\[
\text{ppm H}_2S^{PB} = \text{ppm H}_2S^{RB} \times [1 - \text{RH}_2S] 
\] (5-24)

Where:

ppm H₂Sₚₘₜ = H₂S in purified biogas, ppm
ppm H₂Sᵣₚₘₜ = H₂S in raw biogas, ppm
An assumed value of ppm H₂Sᵣₚₘₜ = 300 ppm is used.
\text{RH}_2S = \text{H}_2S \text{ Pretreatment Efficiency, } \%

5.2.3.4 Daily Purified Biogas Produced

\[
\text{Q}_{PB} = (\text{Q}_{RB} - \text{Q}_{RBE}) - \text{Q}_{CO_2r} - \text{Q}_{H2Sr}, \text{ which reduces to:}
\]

\[
\text{Q}_{PB} = (\text{Q}_{RB} - \text{Q}_{RBE}) \times [1 - \%\text{CO}_2^{RB} \times \text{RCO}_2 - \text{ppmH}_2S^{RB} \times \text{RH}_2S / Z_{volume}] 
\] (5-25)

Where:

\text{Q}_{PB} = \text{Daily purified biogas produced, scf/d (m}^3\text{/d)}
\text{Q}_{RB} = \text{Daily raw biogas produced, scf/d (m}^3\text{/d)}
\text{Q}_{RBE} = \text{Daily raw biogas for electrical energy production, scf/d (m}^3\text{/d)}
\%\text{CO}_2^{RB} = \text{Percentage of CO}_2 \text{ in raw biogas, } \%
\text{RCO}_2 = \text{CO}_2 \text{ Pretreatment Efficiency, } \%
\text{ppm H}_2S^{RB} = \text{H}_2S \text{ in raw biogas, ppm}
An assumed value of ppm H₂Sᵣₚₘₜ = 300 ppm is used.
\[ \text{RH}_2\text{S} = \text{H}_2\text{S} \] Pretreatment Efficiency, \%  
\[ Z_{\text{volume}} = \text{Conversion factor for volume, } 10^6\text{mL/m}^3 \]

5.2.3.5 Percentage of CH\(_4\) in Purified Biogas
\[ \%\text{CH}_4\text{PB} = \%\text{CH}_4\text{RB} \times \frac{Q\text{RB}}{Q\text{PB}} \] (5-26)

Where:
\%\text{CH}_4\text{PB} = \text{Percentage of methane in purified biogas, \%}
\%\text{CH}_4\text{RB} = \text{Percentage of CH}_4 \text{ in raw biogas, \%}
An assumed value of \%\text{CH}_4\text{RB} = 60\% is used unless otherwise specified by the user.
\[ Q_{\text{RB}} = \text{Daily raw biogas produced, scf/d (m}^3/d) \]
\[ Q_{\text{PB}} = \text{Daily purified biogas produced, scf/d (m}^3/d) \]

5.2.3.6 Energy Recovered from Biogas
Based on previous decisions made regarding ERSs in use, and recovered energy efficiencies, the energy recovered by processes declared in use is next computed.

5.2.3.7 Amount of Daily Purified Biogas for Energy

For electrical energy production
\[ Q_{\text{PBEE}} = \frac{Z_{\text{time}} \times C_{\text{ERS}} \times \%\text{CH}_4\text{PB} \times \text{ECE} \times \text{EC}_{\text{CH}_4} \times Z_{\text{energy}}}{Z_{\text{time}} \times \%\text{CH}_4\text{RB} \times \text{ECE} \times \text{EC}_{\text{CH}_4} \times Z_{\text{energy}}} \] (5-27)

Where:
\[ Q_{\text{PBEE}} = \text{Amount of daily purified biogas for electrical energy production, scf/d (m}^3/d) \]
\[ Z_{\text{time}} = \text{Conversion factor for time, 86400s/d} \]
\[ C_{\text{ERS}} = \text{Capacity of energy recovery system, kW} \]
\%\text{CH}_4\text{PB} = \text{Percentage of CH}_4 \text{ in purified biogas, \%}
\text{ECE} = \text{Energy recovery efficiency, \%}
\text{EC}_{\text{CH}_4} = \text{Energy content of methane, Btu/scf (MJ/m}^3) \]
An assumed value of \text{EC}_{\text{CH}_4} = 961.5\text{Btu/scf (35.8 MJ/m}^3) \text{ is used.}
\[ Z_{\text{energy}} = \text{Conversion factor for energy, 1.054kJ/Btu (1000kJ/MJ)} \]

For mechanical energy production
\[ Q_{\text{PBME}} = \frac{Z_{\text{time}} \times C_{\text{ERS}} \times \%\text{CH}_4\text{PB} \times \text{ECE} \times \text{EC}_{\text{CH}_4} \times Z_{\text{energy}}}{Z_{\text{time}} \times \%\text{CH}_4\text{RB} \times \text{ECE} \times \text{EC}_{\text{CH}_4} \times Z_{\text{energy}}} \] (5-28)

Where:
\[ Q_{\text{PBME}} = \text{Amount of daily purified biogas for mechanical energy production, scf/d (m}^3/d) \]
Z_{time} = Conversion factor for time, 86,400s/d
C_{ERS} = Capacity of energy recovery of direct drive, hp (kW)
%CH_{4PB} = Percentage of CH_{4} in purified biogas, %
ECE = Energy recovery efficiency, %
EC_{CH_{4}} = Energy content of CH_{4}, Btu/scf (MJ/m^{3})
An assumed value of EC_{CH_{4}} = 961.5Btu/scf (EC_{CH_{4}} = 35.8 MJ/m^{3}) is used.
Z_{energy} = Conversion factor for energy, hp / 61091Btu/d (1000kJ/MJ)

\textit{Amount of daily purified biogas for energy production}
Q_{PBE} = Q_{PBEE} + Q_{PBME} \quad (5-29)

Where:
Q_{PBE} = Amount of daily purified biogas for energy production, scf/d (m^{3}/d)
Q_{PBEE} = Amount of daily purified biogas for electrical energy production, scf/d (m^{3}/d)
Q_{PBEME} = Amount of daily purified biogas for mechanical energy production, scf/d (m^{3}/d)

\textit{5.2.3.8 Amount of Daily Purified Biogas Not Used}
Q_{PBN} = Q_{PB} - Q_{PBE} \quad (5-30)

Where:
Q_{PBN} = Amount of daily purified biogas not used, scf/d (m^{3}/d)
Q_{PB} = Daily purified biogas produced, scf/d (m^{3}/d)
Q_{PBE} = Amount of daily purified biogas for energy production, scf/d (m^{3}/d)

\textit{5.2.4 Energy Recovery System}
The incoming digester gas flow rate is an operational variable because the facility’s management must decide the amount of digester gas that is going to be used, for example, for building space heating.

A general schematic depiction of digester gas energy recovery is presented in Figure 5-2. The technical variables for energy recovery from digester gas are presented in Table 5-5. Gas flowing into the ERS can be digester gas, natural gas, or a mixture of the two. Because management makes the choice, supplementing of natural gas is an operational variable. If the feed gas were a mixture of digester gas and natural gas, its energy content would be the sum of the digester gas energy content and that of the natural gas. Also, because plant operators can vary the power output of their ERS, the operating efficiency of the recovery system becomes another operating variable.
Where:

\[ \text{Gas}_T = \text{Gas flow rate}, \text{ scf/d (m}^3\text{/d)} \]
\[ E_{\text{thermal}} = \text{Thermal energy, Btu/d (MJ/d)} \]
\[ E_{\text{electrical}} = \text{Electrical energy, kWh/d (MJ/d)} \]
\[ E_{\text{mechanical}} = \text{Mechanical energy, hp (MJ/d)} \]
\[ E_{\text{useful}} = \text{Useful energy, Btu/d (MJ/d)} \]
\[ E_{\text{non-useful}} = \text{Non-useful energy, Btu/d (MJ/d)} \]
\[ \text{EM} = \text{emissions, lb/d (kg/d)} \]

Figure 5-2. General Schematic of Energy Recovery System.

<table>
<thead>
<tr>
<th>Input variables</th>
<th>Operating variables</th>
<th>Intermediate variables</th>
<th>Output variables</th>
</tr>
</thead>
<tbody>
<tr>
<td>Digester gas flow rate scf/d (m³/d)</td>
<td>Natural gas flow rate scf/d (m³/d)</td>
<td>Total gas flow rate scf/d (m³/d)</td>
<td>Thermal energy Btu/d (MJ/d)</td>
</tr>
<tr>
<td>% Methane</td>
<td>Operating efficiency</td>
<td>Gas energy content Btu/scf (MJ/m³)</td>
<td>Electrical energy kWh/d (MJ/d)</td>
</tr>
<tr>
<td>Energy recovery system</td>
<td></td>
<td>% Thermal Energy Recovery efficiency</td>
<td>Mechanical energy, hp (MJ/d)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>% Electrical Energy Recovery efficiency</td>
<td>Non-useful energy Btu/d (MJ/d)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>% Mechanical Energy Recovery efficiency</td>
<td></td>
</tr>
</tbody>
</table>

The form and the conversion efficiency of useful energy depend on the facility’s chosen ERS. Boilers can only convert the incoming gas mix into thermal energy, whereas gas engines, microturbines, and fuel cells are capable to simultaneously produce electrical and thermal energy. As opposed to gas engine generators, direct drives produce mechanical instead of electrical energy. Consequently, the amount of thermal, electrical, and mechanical energy produced is reliant on the proportion of the gas mix and its energy content, the energy recovery efficiencies as well as the operating efficiency of the energy recovery system. Non-useful energy is energy lost within the ERS. Emission rates of combustion gases such as CO₂, NOₓ, and SOₓ are dependent on the chosen ERS.
5.2.4.1 Thermal Energy

Thermal Energy Recovered

\[ \text{TER} = \text{ECE}_{\text{Ther}} \times \text{Q}_{\text{RBTE}} \times \%\text{CH}_{4\text{RB}} \times \text{EC}_{\text{CH}_4} \]  \hspace{1cm} (5-31)

Where:

- \( \text{TER} \) = Thermal energy recovered, Btu/d (MJ/d)
- \( \text{ECE}_{\text{Ther}} \) = Thermal energy recovery efficiency, %
- \( \text{Q}_{\text{RBTE}} \) = Amount of raw biogas for thermal energy, scf/d (m³/d)
- \( \%\text{CH}_{4\text{RB}} \) = Percentage of \( \text{CH}_4 \) in raw biogas, %
  
  Assumed value of \( \%\text{CH}_{4\text{RB}} \) = 60% is used.
- \( \text{EC}_{\text{CH}_4} \) = Energy content of \( \text{CH}_4 \), Btu/scf (MJ/m³)
  
  Assumed value of \( \text{EC}_{\text{CH}_4} = 961.5 \) Btu/scf (35.8 MJ/m³) is used.

Net Energy (Thermal) Recovered

The net energy recovered at the WWTP site is only thermal energy, and is that which can be recovered for heating of the digestion tank or the digester feed sludge.

\[ \text{NTER} = \text{TER} - \frac{\text{EDH}}{\text{Z}_{\text{energy}}} \]  \hspace{1cm} (5-32)

Where:

- \( \text{NTER} \) = Net thermal energy recovered, Btu/d (MJ/d)
- \( \text{TER} \) = Thermal energy recovered, Btu/d (MJ/d)
- \( \text{EDH} \) = Energy needed for digester heating, Btu/d (W)
- \( \text{Z}_{\text{energy}} \) = Conversion factor for energy, 1Btu/d/Btu/d (11.574W/MJ/d)

5.2.4.2 Mechanical Energy

Mechanical Energy Recovered

\[ \text{MER} = \left( \frac{\text{ECE}_{\text{Mech}} \times \text{Q}_{\text{PBME}} \times \%\text{CH}_{4\text{PB}} \times \text{EC}_{\text{CH}_4}}{\text{Z}_{\text{energy}}} \right) \]  \hspace{1cm} (5-33)

Where:

- \( \text{MER} \) = Mechanical energy recovered, hp (MJ/d)
- \( \text{ECE}_{\text{Mech}} \) = Mechanical energy recovery efficiency, %
- \( \text{Q}_{\text{PBME}} \) = Amount of purified biogas for mechanical energy production, scf/d (m³/d)
- \( \%\text{CH}_{4\text{PB}} \) = Percentage of \( \text{CH}_4 \) in purified biogas, %
- \( \text{EC}_{\text{CH}_4} \) = Energy content of \( \text{CH}_4 \), Btu/scf (MJ/m³)
  
  An assumed value of \( \text{EC}_{\text{CH}_4} = 961.5 \) Btu/scf (35.8 MJ/m³) is used.
- \( \text{Z}_{\text{energy}} \) = Conversion factor for energy, 61,084 Btu/hp-d (1MJ/1MJ)
**Net Mechanical Energy Recovered**
The net mechanical energy recovered is identical to the total mechanical energy recovered.

**5.2.4.3 Net Electrical Energy Recovered**
The net electrical energy recovered is identical to the total electrical energy recovered.

**5.2.4.4 Emissions from Energy Recovery Systems**
Emission factors for various biogas ERSSs have been located, mostly from US EPA sources, for NOx, SO2, CO, TPM, non-methane volatile organic compounds (NMVOC) and GHG. Canadian emission data from electricity generation plants and the distribution of electricity generation methods for the Provinces in Canada have also been obtained. The emission factors were expressed as mass units of contaminants per quantity of energy produced, as provided in Table 5-6. Air emission loadings of these contaminants may be estimated for the different energy recovery technologies. The emission factor table indicates the engine-generators have the highest factors for CO, NMVOC and NOx, while microturbines and Stirling engines have the highest emission factor for TPM. Depending on the size of the ERS unit geographical location and jurisdictional air regulations, the decision to adopt a technology may be impacted by the mass emission rates of contaminants from the recovery units.

<table>
<thead>
<tr>
<th>Energy Recovery Technology</th>
<th>NOx</th>
<th>NMVOC</th>
<th>TPM</th>
<th>CO</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>lb/MM BtuBG</td>
<td>g/MJBG</td>
<td>lb/MM BtuBG</td>
<td>g/MJBG</td>
<td>lb/MM BtuBG</td>
</tr>
<tr>
<td>Boiler</td>
<td>0.065</td>
<td>0.028</td>
<td>0.009</td>
<td>0.004</td>
<td>0.003</td>
</tr>
<tr>
<td>Engine-Generator</td>
<td>1.255</td>
<td>0.540</td>
<td>0.033</td>
<td>0.014</td>
<td>0.006</td>
</tr>
<tr>
<td>Steam Turbine</td>
<td>0.160</td>
<td>0.069</td>
<td>0.006</td>
<td>0.003</td>
<td>0.012</td>
</tr>
<tr>
<td>Gas Turbine</td>
<td>0.160</td>
<td>0.069</td>
<td>0.006</td>
<td>0.003</td>
<td>0.012</td>
</tr>
<tr>
<td>Microturbine</td>
<td>0.060</td>
<td>0.026</td>
<td>0.275</td>
<td>0.118</td>
<td>0.033</td>
</tr>
<tr>
<td>Molten Carbonate Fuel Cell</td>
<td>0.008</td>
<td>0.003</td>
<td>0.001</td>
<td>0.001</td>
<td>0.000</td>
</tr>
<tr>
<td>Phosphoric Acid Fuel Cell</td>
<td>0.001</td>
<td>0.001</td>
<td>0.002</td>
<td>0.001</td>
<td>0.000</td>
</tr>
<tr>
<td>Stirling Engine</td>
<td>0.041</td>
<td>0.018</td>
<td>0.008</td>
<td>0.004</td>
<td>0.114</td>
</tr>
<tr>
<td>Direct Drive</td>
<td>1.255</td>
<td>0.540</td>
<td>0.033</td>
<td>0.014</td>
<td>0.006</td>
</tr>
<tr>
<td>Combine Cycle</td>
<td>0.160</td>
<td>0.069</td>
<td>0.006</td>
<td>0.003</td>
<td>0.012</td>
</tr>
</tbody>
</table>

**Onsite CO2 Emissions**
CO2 emissions for the various ERSs are based on a methane carbon mass balance. The equations presented next are those for a boiler, but the equations are similarly applied to the other ERSs.
\[ E_{CO2} = \frac{Z_{time} \cdot Q_{RBTE} \cdot [D_{CO2} \cdot \%CO2_{PB} + Z_{CO2/CH4} \cdot D_{CH4} \cdot \%CH4_{PB} \cdot EC_{boiler}]}{Z_{weight}} \quad (5-34) \]

Where:

- \( E_{CO2} \) = Emission of CO\(_2\) from the boiler, ton\(CO2/y\) (tonne\(CO2/y\))
- \( Z_{time} \) = Conversion factor time, 365d/y
- \( Q_{RBTE} \) = Daily raw biogas for thermal energy production, scf/d (\(m^3/d\))
- \( \%CO2_{RB} \) = Percentage of CO\(_2\) in raw biogas, %
  
  \( \text{Assumed value of } \%CO2_{RB} = 40\% \text{ is used.} \)
- \( Z_{CO2/CH4} \) = Conversion factor, 2.75lb\(CO2/lbCH4\) (2.75kg\(CO2/kgCH4\))
- \( \%CH4_{RB} \) = Percentage of CH\(_4\) in raw biogas, %
  
  \( \text{Assumed value of } \%CH4_{RB} = 60\% \text{ is used.} \)
- \( D_{CO2} \) = Density of CO\(_2\)
  
  An assumed value of \( D_{CO2} = 0.123\text{lb/scf (1.964 kg/m}^3) \) is used.
- \( D_{CH4} \) = Density of CH\(_4\)
  
  An assumed value of \( D_{CH4} = 0.045\text{lb/scf (0.7158 kg/m}^3) \) is used.
- \( EC_{boiler} \) = Combustion efficiency of the boiler (as decimal fraction)
- \( Z_{weight} \) = Conversion factor for weight, 2000lb/ton (1000kg/tonne)

**Onsite CH\(_4\) Emissions**

\( \text{CH}_4 \) emissions for the various ERSs are based on the efficiency of methane combustion by the ERS. The equations presented next are those for a boiler, but the equations are similarly applied to the other ERSs.

\[ E_{CH4} = D_{CH4} \cdot \%CH4_{PB} \cdot Q_{RBTE} \cdot (1 - EC_{boiler}) \cdot Z_{time}/Z_{weight} \quad (5-35) \]

Where:

- \( E_{CH4} \) = Emission of CH\(_4\) from the boiler, ton\(CH4/y\) (tonne\(CH4/y\))
- \( Q_{RBTE} \) = Daily raw biogas for thermal energy production, scf/d (\(m^3/d\))
- \( \%CH4_{RB} \) = Percentage of CH\(_4\) in raw biogas, %
  
  \( \text{An assumed value of } \%CH4_{RB} = 60\% \text{ is used.} \)
- \( D_{CH4} \) = Density of CH\(_4\)
  
  An assumed value of \( D_{CH4} = 0.045\text{lb/scf (0.7158 kg/m}^3) \) is used.
- \( EC_{boiler} \) = Combustion efficiency of the boiler (as decimal fraction)
- \( Z_{time} \) = Conversion factor time, 365d/y
- \( Z_{weight} \) = Conversion factor for weight, 2000lb/ton (1000kg/tonne)
**Onsite SO₂ Emissions**

SO₂ emissions for the various ERSs are based on a hydrogen sulfide - sulfur mass balance. The equations presented next are those for a boiler, but the equations are similarly applied to the other ERSs. Emissions of SO₂ from fuel cells are considered to be negligible.

\[ E_{SO2} = \text{ppm } H2S_{RB} \times Q_{RBTE} \times M_{SO2} \times Z_{time} / (n \times Z_{volume} \times Z_{weight}) \]  
(5-36)

Where:

- \( E_{SO2} \) = Emission of SO₂ from boiler, tonSO₂/y (tonneSO₂/y)
- ppm \( H2S_{RB} \) = H₂S in raw biogas, ppm
  - An assumed value of ppm \( H2S_{RB} \) = 300 ppm is used.
- \( M_{SO2} \) = Molecular weight of SO₂, 64lb/mol (64g/mol)
- \( Z_{time} \) = Conversion factor time, 365d/y
- \( n \) = Number of moles per unit volume, 0.8638scf/mol (24.46L/mol)
- \( Z_{volume} \) = Conversion factor for volume, 1 scf/scf (1000L/m³)
- \( Z_{weight} \) = Conversion factor for weight, 2000lb/ton (1000kg/tonne)
- \( Q_{RBTE} \) = Daily raw biogas for thermal energy production, scf/d (m³/d)

**5.2.4.5 Avoided Offsite Emissions**

Finally, avoided offsite emissions are calculated as a credit for the WWTPs. The calculations are provided to account for the fact the energy produced and utilized on-site offsets energy produced elsewhere by means of thermal, hydro, nuclear, or other generating means. Emission rates for electricity generation in each state or province will be used to estimate the avoided offsite emissions.

**Avoided Offsite CO₂ Emissions**

\[ E_{offset CO2} = Q_{bioelec-s} \times epk_{CO2,J} \]  
(5-37)

Where:

- \( E_{offset CO2} \) = Offset emission of CO₂, tonCO₂/y (tonneCO₂/y)
- \( Q_{bioelec-s} \) = Annual produced and plant-consumed electricity, kWh/y
- \( epk_{CO2,J} \) = Amount of CO₂ generated per kWh of electricity produced in a Jurisdiction, tonCO₂/kWh (tonneCO₂/kWh)

**Avoided Offsite SO₂ Emissions**

\[ E_{offset SO2} = Q_{bioelec-s} \times epk_{SO2,J} \]  
(5-38)

Where:

- \( E_{offset SO2} \) = Offset emission of SO₂, tonSO₂/y (tonneSO₂/y)
\[ Q_{\text{bioelec-s}} = \text{Annual produced and plant-consumed electricity, kWh/y} \]
\[ \text{Epk}_{\text{SO2,J}} = \text{Amount of SO2 generated per kWh of electricity produced in a Jurisdiction, tonSO2/kWh (tonneSO2/kWh)} \]

**Avoided Offsite NOX Emissions**
\[ E_{\text{offset NOX}} = Q_{\text{bioelec-s}} \times \text{epk}_{\text{NOX,J}} \]  
(5-39)

Where:
\[ E_{\text{offset NOX}} = \text{Offset emission of NOX, tonNOx/y (tonneNOx/y)} \]
\[ Q_{\text{bioelec-s}} = \text{Annual produced and plant-consumed electricity, kWh/y} \]
\[ \text{Epk}_{\text{NOX,J}} = \text{Amount of NOX generated per kWh of electricity produced in a Jurisdiction, tonNOx/kWh (tonneNOx/kWh)} \]

**Avoided Offsite VOC Emissions**
\[ E_{\text{offset VOC}} = Q_{\text{bioelec-s}} \times \text{epk}_{\text{VOC,J}} \]  
(5-40)

Where:
\[ E_{\text{offset VOC}} = \text{Offset emission of VOC, tonVOC/y (tonneVOC/y)} \]
\[ Q_{\text{bioelec-s}} = \text{Annual produced and plant-consumed electricity, kWh/y} \]
\[ \text{Epk}_{\text{VOC,J}} = \text{Amount of VOC generated per kWh of electricity produced in a Jurisdiction, tonVOC/kWh (tonneVOC/kWh)} \]

**Avoided Offsite PM10 Emissions**
\[ E_{\text{offset PM10}} = Q_{\text{bioelec-s}} \times \text{epk}_{\text{PM10,J}} \]  
(5-41)

Where:
\[ E_{\text{offset PM10}} = \text{Offset emission of PM10, tonPM10/y (tonnePM10/y)} \]
\[ Q_{\text{bioelec-s}} = \text{Annual produced and plant-consumed electricity, kWh/y} \]
\[ \text{Epk}_{\text{PM10,J}} = \text{Amount of PM10 generated per kWh of electricity produced in a Jurisdiction, tonPM10/kWh (tonnePM10/kWh)} \]

**Avoided Offsite CO Emissions**
\[ E_{\text{offset CO}} = Q_{\text{bioelec-s}} \times \text{epk}_{\text{CO,J}} \]  
(5-42)

Where:
\[ E_{\text{offset CO}} = \text{Offset emission of CO, tonCO/y (tonneCO/y)} \]
\[ Q_{\text{bioelec-s}} = \text{Annual produced and plant-consumed electricity, kWh/y} \]
\( E_{p,CO, \text{J}} \) = Amount of CO generated per kWh of electricity produced in a Jurisdiction, ton\( _{\text{CO}}/\text{kWh} \) (tonne\( _{\text{CO}}/\text{kWh} \)

5.3 Economic Functions for the LCAMER Spreadsheet Model

All estimates and statistically valid relations between unit costs and applicable variables for the LCC method are presented in this section. These results were developed from the database, literature review and vendor quotations, and are presented for general variables, anaerobic digestion parameters, GPSs parameters, and ERSs parameters. The economic functions following are based on US dollars and are current as of June 2005.

The economic functions developed in this project required that certain values be assumed. The default values used and recommended in the study are as presented in Table 5-7.

<table>
<thead>
<tr>
<th>Item</th>
<th>Unit</th>
<th>Suggested Default Values (Ranges)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Planning period</td>
<td>Year</td>
<td>20</td>
</tr>
<tr>
<td>Discount rate</td>
<td>%</td>
<td>6 (3.00-9.00)</td>
</tr>
<tr>
<td>Electricity price</td>
<td>$/kWh</td>
<td>0.09 (0.02-0.24)</td>
</tr>
<tr>
<td>Natural gas price</td>
<td>$/scf ($/m³)</td>
<td>0.007 [0.001-0.023] (0.25 [0.02-0.82])</td>
</tr>
<tr>
<td>Diesel price</td>
<td>$/gal ($/L)</td>
<td>1.51 [0.91-2.95] (0.4 [0.24-0.78])</td>
</tr>
<tr>
<td>Construction element installation cost factor</td>
<td>%</td>
<td>100</td>
</tr>
<tr>
<td>Contractor's overhead and profit cost factor</td>
<td>%</td>
<td>15</td>
</tr>
<tr>
<td>Engineering</td>
<td>%</td>
<td>15</td>
</tr>
<tr>
<td>Contingency allowance</td>
<td>%</td>
<td>15</td>
</tr>
<tr>
<td>Service Life</td>
<td>Structure</td>
<td>Year</td>
</tr>
<tr>
<td></td>
<td>Gas Pretreatment Technology</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Boiler</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Gas Engine Cogenerator</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Steam Turbine</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Gas Turbine</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Direct Drive</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Microturbine</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Fuel Cell</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Stirling Engine</td>
<td></td>
</tr>
</tbody>
</table>

5.3.1 Capital Costs

5.3.1.1 Anaerobic Digester

*Anaerobic Digester Unit Capital Cost*

The unit capital cost functions are empirical relationships derived from project survey data. Details of the derivations are found in Liu (2005).
Conventional concrete wall

\[ \text{UC}_{\text{AD}} = A \times V_T^{-0.322} \]  \hspace{1cm} (5-43)
\[ R^2 = 0.759 \]

Egg-shaped steel wall

\[ \text{UC}_{\text{AD}} = B \times V_T^{-0.1514} \]  \hspace{1cm} (5-44)
\[ R^2 = 0.869 \]

Where:
- \( \text{UC}_{\text{AD}} \) = Unit capital cost of anaerobic digester, $/gal ($/m³)
- \( V_T \) = Total volume of digester, gal (m³)
  - \( A = 17593 \) and \( B = 5179.4 \) if \( V_T \) in m³
  - \( A = 401.22 \) and \( B = 45.62 \) if \( V_T \) in gal

Egg-shaped concrete wall

Insufficient survey data were available to derive an empirical relationship. The unit cost from the limited data of one facility was estimated as $7.00/gal ($1,850/m³).

**Anaerobic Digester Capital Cost**

\[ \text{IC}_{\text{AD}} = \text{UC}_{\text{AD}} \times V_T \]  \hspace{1cm} (5-45)

Where:
- \( \text{IC}_{\text{AD}} \) = Initial capital cost of anaerobic digester, $
- \( \text{UC}_{\text{AD}} \) = Unit capital cost of anaerobic digester, $/gal ($/m³)
- \( V_T \) = Total volume of digester including gas headspace, gal (m³)

### 5.3.1.2 Energy Recovery Systems

**Unit Capital Cost of ERS**

The unit capital costs for the ERSs are based on project survey and literature results.

The unit cost factors derived, expressed in units of $/kW, are provided in Table 5-8. An empirical relationship was developed for only one ERS, namely the microturbines of capacity in the 30 – 120 kW range. For that type of system only the unit cost is expressed as:

\[ \text{UC}_{\text{MT}} = 14.54 \times \text{CPERS} + 4851 \]  \hspace{1cm} (5-46)

Where:
- \( \text{UC}_{\text{MT}} \) = Unit cost of microturbine in 30 – 120 kW capacity range, $/kW
- \( \text{CPERS} \) = Capacity of ERS, kW
Table 5-8. Unit Capital Cost Factors for Energy Recovery Systems.

<table>
<thead>
<tr>
<th>Technology</th>
<th>Unit Capital Costs</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Total Capacity¹</td>
</tr>
<tr>
<td>Boiler</td>
<td>2,944 (10,000 MBtu/h)</td>
</tr>
<tr>
<td>Gas Engine Co-Generator</td>
<td>&lt;1,000</td>
</tr>
<tr>
<td></td>
<td>1,000-3,000</td>
</tr>
<tr>
<td></td>
<td>&gt;3,000</td>
</tr>
<tr>
<td>Steam Turbine</td>
<td>8,000-20,000</td>
</tr>
<tr>
<td>Gas Turbine</td>
<td>1,000-10,000</td>
</tr>
<tr>
<td>Combined Cycle</td>
<td>35,200</td>
</tr>
<tr>
<td>Direct Drive</td>
<td>100 (134.05hp)</td>
</tr>
<tr>
<td>Microturbine</td>
<td>30-120</td>
</tr>
<tr>
<td></td>
<td>120-300</td>
</tr>
<tr>
<td></td>
<td>≥300</td>
</tr>
<tr>
<td>Fuel Cell</td>
<td>Molten Carbon DFC300</td>
</tr>
<tr>
<td></td>
<td>Phosphoric Acid PC25</td>
</tr>
<tr>
<td>Stirling Engine</td>
<td>55</td>
</tr>
</tbody>
</table>

¹ In units of kW unless otherwise specified
² Where x is total capacity of microturbine

**Capital Cost of ERS**

For all ERSs, the initial capital cost is calculated as:

\[ IC_{ERS} = UC_{ERS} \times C_{PERS} \]  

Where:

- \( IC_{ERS} \) = Initial capital cost of ERS, $
- \( UC_{ERS} \) = Unit capital cost of ERS, $/ Btu/h or $/hp ($/kW)
- \( C_{PERS} \) = Capacity of ERS, Btu/h or hp (kW)

**5.3.1.3 Gas Purification Systems**

Four types of substances have been identified for removal from the raw biogas. These include moisture, \( H_2S \), siloxanes and \( CO_2 \). The capital cost functions for the types of gas purification systems follow.

**Unit Capital Cost of GPS**

Unit capital cost factors derived from project survey data and literature are provided in Table 5-9. Because of the basis from which the unit cost factors were derived, the units in which the factors are expressed are not consistent, but are mostly based on digester gas flow rate. Unit capital costs could not be derived for the LO-CAT system for \( H_2S \) removal.
Table 5-9. Unit Capital Cost Factors for Gas Pretreatment Systems.

<table>
<thead>
<tr>
<th>Technology</th>
<th>Based on</th>
<th>Total Capacity</th>
<th>Default value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Iron Salts Addition</td>
<td>Raw Wastewater (design capacity)</td>
<td>167MGD</td>
<td>292$/MGD</td>
</tr>
<tr>
<td>AD Sludge Feed Line</td>
<td>3.69 ton/d</td>
<td>3,870</td>
<td>2.49</td>
</tr>
<tr>
<td>Iron Sponge</td>
<td>Digester Gas</td>
<td>136,668</td>
<td>19,313$/ton/d</td>
</tr>
<tr>
<td>Sufa Treat</td>
<td>Digester Gas</td>
<td>136,668</td>
<td>136,668</td>
</tr>
<tr>
<td>Water Scrubber</td>
<td>Digester Gas</td>
<td>224,990</td>
<td>224,990</td>
</tr>
<tr>
<td>Apollo</td>
<td>Digester Gas</td>
<td>432,322</td>
<td>432,322</td>
</tr>
<tr>
<td>LO-CAT</td>
<td>Digester Gas</td>
<td>N/A</td>
<td>High capital investment</td>
</tr>
<tr>
<td>Activated Carbon Adsorption</td>
<td>Digester Gas</td>
<td>1,589,000</td>
<td>1,589,000</td>
</tr>
</tbody>
</table>

1Unless otherwise stated

**Unit capital cost of iron salts addition**
Addition to raw wastewater

\[ \text{UC}_{\text{GPS}} = \frac{292}{\text{MGD}_{\text{raw wastewater}}} \left( \frac{0.077}{\text{m}^3_{\text{raw wastewater}}} \right) \]

Addition to digester sludge feed line

\[ \text{UC}_{\text{GPS}} = \frac{19,313}{\text{ton}_{\text{VS}}} \left( \frac{21,289}{\text{tonne}_{\text{VS}}} \right) \]

**Unit capital cost of other gas pretreatment systems**
The unit capital cost for all other GPS is expressed in units of $/m3/d of digester gas flow.

**Capital Cost of GPS**

**Initial capital cost of iron salts addition**

Iron salts added to raw wastewater

\[ \text{IC}_{\text{GPS}} = \text{UC}_{\text{GPS}} \times Q_W \quad (5-48) \]

Where:

\[ \text{IC}_{\text{GPS}} = \text{Initial cost of GPS (iron salts), } \$
\]

\[ \text{UC}_{\text{GPS}} = \text{Unit cost of GPS (iron salts), } \frac{\$/\text{MGD}_{\text{raw wastewater}}}{\text{$/m}^3_{\text{raw wastewater}}} \]

\[ Q_W = \text{Daily wastewater flow, MGD (m}^3/\text{d)} \]
Iron salts added to digester feed sludge line

\[ \text{IC}_{\text{GPS}} = \text{UC}_{\text{GPS}} \times \text{DT}_S \]  
(5-49)

Where:

\( \text{IC}_{\text{GPS}} \) = Initial capital cost of GPS (iron salts), $
\( \text{C}_{\text{GPS}} \) = Unit cost of GPS (iron salts), $/ton_{VS/d} ($/tonne_{VS/d})
\( \text{DT}_S \) = Digester feed solids mass, ton/d (tonne_{VS/d})

**Initial capital cost of other GPS**

For other GPSs,

\[ \text{IC}_{\text{GPS}} = \text{UC}_{\text{GPS}} \times \text{Q}_{\text{PB}} \]  
(5-50)

Where:

\( \text{IC}_{\text{GPS}} \) = Initial capital cost of GPS, $
\( \text{UC}_{\text{GPS}} \) = Unit capital cost of GPS, $/scf_{PB/d} ($/m^3_{PB/d})
\( \text{Q}_{\text{PB}} \) = Daily raw biogas for pretreatment, scf/d (m^3/d)

**Replacement Cost of Gas Pretreatment System**

Although the overall replacement period for the digestion process and ERS may be specified for a longer period (e.g. 20 years), within that time frame it may be necessary to replace sub-components more regularly, such as the GPSs. Knowing the number of time of replacement within the lifetime allow the calculation of the replacement cost, expressed in current dollars. The cost for replacing the GPS within the planning period is given as:

\[ \text{RC}_{\text{GPS}} = \text{IC}_{\text{GPS}} \times \left[ \frac{1}{(1+i)^{RS}} + \frac{1}{(1+i)^{2RS}} + \ldots + \frac{1}{(1+i)^{NR \times RS}} \right] \]  
(5-51)

Where:

\( \text{RC}_{\text{GPS}} \) = Replacement cost of GPS, $
\( \text{IC}_{\text{GPS}} \) = Initial capital cost of GPS, $
\( \text{RS} \) = Replacement schedule of GPS, y
\( \text{NR} \) = Round number of replacement of GPS during the planning period
\( i \) = Discount rate, %

A default value of \( i = 6\% \) was assumed.

**5.3.1.4 Net Capital Cost**

The net capital cost takes into consideration the capital costs for the anaerobic digestion tanks, ERSs and capital and replacement costs for the GPSs. Then, should any capital grant be
available to help to offsite the capital costs, the value of the grant may be deducted from the capital and replacement costs.

The net capital cost is given as:

\[
CC = IC_{AD} + IC_{ERS} + IC_{GPS} + RCGPS - G 
\]

Where:
- \(CC\) = Net capital cost, $
- \(IC_{AD}\) = Initial capital cost of anaerobic digester, $
- \(IC_{ERS}\) = Initial capital cost of ERS, $
- \(IC_{GPS}\) = Initial capital cost of GPS, $
- \(RC_{GPS}\) = Replacement cost of GPS, $
- \(G\) = Grant received, $

### 5.3.2 Surplus Biogas-Generated Electricity (Bioelectricity)

When a treatment facility is generating electricity from the digester biogas, there is opportunity both, to use the bioelectricity produced on-site to off-set purchased electricity from a utility or to deliver the bioelectricity offsite for sale. The modeling equations account for the bioelectricity produced and consumed on-site, the need for additional purchased electricity from offsite, and a decision whether to use any of the bioelectricity for offsite delivery. It is possible to calculate an estimated demand for offsite energy if the default option is initially selected. The model uses a relationship for treatment facility off-site electricity demand based on population served by the treatment plant. Otherwise, the user can input the offsite electricity demand for the year. The user also has the option of using the value of model-calculated onsite bioelectricity produced and consumed, or of using a default value based on a percentage of offsite electricity demand.

A WWTP may be generating more bioelectricity than it uses, but yet may be importing purchased electricity while at the same time reserving some bioelectricity for offsite sale (perhaps to fulfill a contractual obligation). The model calculates the plant’s offsite electricity demand, and gives the user the option of transferring offsite shipment of bioelectricity back to onsite use to offset the purchased electricity. The function used is shown in Equation 5-53.

\[
Q_{surbioelec} = OPBE * Z_{time} - AelD 
\]

Where:
- \(Q_{surbioelec}\) = Electricity (bio) surplus, kWh/y
- \(OPBE\) = Onsite produced bioelectricity, kWh/d
- \(Z_{time}\) = Conversion factor for time, 365d/y
- \(AelP\) = Annual off-site electricity purchased (off-site electricity demand), kWh/y
5.3.3 Salvage Value of Equipment

This section accounts for the residual value of equipment used for the digester gas production and ERS, once their useful life is over and they are scrapped. Two scenarios are possible. The first is when equipment has a shorter life than the planning period, and so must be replaced more frequently, and the other in which the equipment life is equal to or greater than the planning period.

5.3.3.1 Net Salvage Value for Equipment with Shorter Life than Planning Period

This scenario is applicable to all ERSs considered in the project.

\[
SV_{ERS} = IC_{ERS} \frac{[RS_{ERS} - (N - RS_{ERS} \times NR_{ERS})]}{RS_{ERS}} \times \frac{1}{(1 + i)^N}
\]  

(5-54)

Where:

- \(SV_{ERS}\) = Net salvage value of ERS, $
- IC_{ERS}\) = Initial capital cost of ERS, $
- RS_{ERS}\) = Replacement schedule for ERS, y
- \(NR_{ERS}\) = Round number of replacement of ERS during the planning period
- \(N\) = Planning period, y
- A default value of \(N = 20\) years was assumed.
- \(i\) = Discount rate, %
- A default value of \(i = 6\%\) was assumed.

5.3.3.2 Net Salvage Value for Equipment with Longer/Equal Life to Planning Period

Equipment such as the anaerobic digester should be considered in this application because it has projected 40 years service life. The salvage value of such equipment is a straight-line depreciation is given by:

\[
SV_{LE} = IC \times \frac{[RS - N]}{RS}
\]  

(5-55)

Where:

- \(SV_{LE}\) = Net salvage value of equipment with service life longer than planning period, $
- IC\) = Initial capital cost of equipment, $
- RS\) = Replacement schedule (service life) for equipment, y
- \(N\) = Planning period, y
- A default value of \(N = 20\) years was assumed.
5.3.3.3 Net Salvage Value

\[ SV_S = SV_{ERS} + SV_{LE} \]  

Where:

\( SV_S \) = Net salvage value total, $
\( SV_{ERS} \) = Net salvage value of ERS, $
\( SV_{LE} \) = Net salvage value of equipment with service life longer or equal to planning period, $

5.3.3.4 Annualized Salvage Value

The annualized salvage value uses the salvage value to convert the net salvage value to an annualized cost.

\[ ASV_S = SV_S \frac{i(1+i)^N}{(1+i)^N-1} \]  

Where:

\( ASV_S \) = Annualized salvage value, $/y
\( SV_S \) = Net salvage value, $
\( N \) = Planning period, y
A value of \( N = 20 \) years was assumed.
\( i \) = Discount rate, %
A value of \( i = 6\% \) was assumed.

5.3.4 Net Annual Revenue

Revenue is generated by potential sales of bioelectricity, recovered heat, or biogas to offsite customers, in addition to revenue from potential GHG emission credits.

\[ AR = P_{bioelec} * Q_{surbioelec} * Z_{time} * (P_{bioheat} * NTER * Z_{energy} + P_{biogas} * Q_{PBN}) + E_{offsetCO2} * cr_{GHG} \]  

Where:

\( AR \) = Net annual revenue, $/y
\( P_{bioelec} \) = Selling price of bio-electricity, $/kWh
\( Q_{surbioelec} \) = Surplus bioelectricity, kWh/y
\( Z_{time} \) = Conversion factor for time, 365d/y
\( P_{bioheat} \) = Selling price of bioheat, $/Btu ($/kJ)
\( NTER \) = Net thermal energy recovered, Btu/d (MJ/d)
\( Z_{energy} \) = Conversion factor for energy, 1Btu/Btu (1000kJ/MJ)
P_{biogas} = Selling price of the biogas, $/scf ($/m^3)
Q_{PBN} = Amount of daily purified biogas not used, scf/d (m^3/d)
E_{offsetCO2} = Offset emission of CO_2, tonCO_2/y (tonneCO_2/y)
cr_{CO2} = GHG credit for reduced CO_2 emissions, $/tonCO_2 ($/tonneCO_2)

5.3.4.1 Net Annual Total Savings of Avoided Offsite Energy Consumption and Revenue

Use of the digester biogas offsets the energy needed to operate the digester, and potentially the rest of treatment plant, depending on the user inputs. The quantities of offsite energy deferred, both electricity and natural gas are calculated. Then, knowing the cost of these purchased commodities, the annual savings from replacing offsite energy with on-site produced heat and bioelectricity can be calculated on an annual basis. Lastly, the total combined savings and revenues are totalled to provide what is termed the net annual total revenue, although properly the value of the offsite energy deferred is a savings rather than a true revenue.

\[
AS = \frac{[%pro-normal \times P_{elec-normal} + %pro-peak \times P_{elec-peak}] \times Q_{bioelec} \times P_{ngas} \times Q_{bioheat}}{EC_{ngas}}
\]  

(5-59)

Where:

\( AS \) = Annual net saving, $/y
\( P_{elec-normal} \) = Normal electricity price paid by the plant, $/kWh
\( P_{elec-peak} \) = Peak electricity price paid by the plant, $/kWh
\( %pro-normal \) = Percentage of annual produced electricity for normal electricity price, %
\( %pro-peak \) = Percentage of annual produced electricity for peak electricity price, %
\( Q_{bioelec-s} \) = Annual produced and plant-consumed electricity, kWh/y
\( P_{ngas} \) = Price of natural gas paid by the plant, $/m^3
\( EC_{ngas} \) = Energy content of natural gas, 1030Btu/scf (38.3 MJ/m^3)
\( Q_{bioheat-s} \) = Annual plant-consumed bioheat, Btu/y (MJ/y)

5.3.4.2 Net Annual Total Revenue

\[
ATR = AR + ASVS
\]

(5-60)

Where:

\( ATR \) = Net annual total revenue ($/y)
\( ASVS \) = Annualized salvage value ($/y)
\( AR \) = Net annual revenue ($/y)
5.3.5 Annual Operating and Maintenance Cost

5.3.5.1 Annual O&M for Anaerobic Digestion Process

*Unit O&M Costs of Digester*

Annual unit O&M costs, expressed as dollars per unit mass of VS loaded to the digester, have been derived from the surveys for the different anaerobic processes, as the number of responses for each type of process permitted. Because of the limited number of egg-shaped digesters in operation, the O&M costs for this type of construction were generally not available. In this study, O&M costs for egg-shaped digesters were estimated by applying a factor of 0.45 to O&M costs for conventionally-shaped digesters. The unit O&M costs for conventional digesters using the different anaerobic processes are provided in Table 5-10.

<table>
<thead>
<tr>
<th>Digestion Process</th>
<th>Unit O&amp;M Anaerobic Digestion Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$/tonVS</td>
</tr>
<tr>
<td>Single stage mesophilic</td>
<td>110.7</td>
</tr>
<tr>
<td>Two-stage mesophilic</td>
<td>88.9</td>
</tr>
<tr>
<td>Acid-Gas Meso/Meso</td>
<td>52.2</td>
</tr>
<tr>
<td>Acid-Gas Meso/Thermo</td>
<td>52.2</td>
</tr>
<tr>
<td>Single-stage thermophilic</td>
<td>102.5</td>
</tr>
<tr>
<td>Two-stage thermophilic</td>
<td>102.5</td>
</tr>
<tr>
<td>Temperature-phased Meso/Thermo</td>
<td>38.8</td>
</tr>
<tr>
<td>Temperature-phased Thermo/Meso</td>
<td>38.8</td>
</tr>
</tbody>
</table>

Note: These costs do not include utilities

Annual O&M Cost of Digester

\[
AOM_{AD} = UOM_{AD} \times DT_S \times Z_{time} \tag{5-61}
\]

Where:

\[AOM_{AD} = \text{Annual O&M cost of digester, } \$/y\]
\[UOM_{AD} = \text{Unit O&M costs of digester, } \$/\text{ton VS (}/$\text{tonne VS)}\]
\[DT_S = \text{Average daily mass of VS loaded to digester, } \text{ton VS/d (}/\text{tonne VS/d)}\]
\[Z_{time} = \text{Conversion factor for time, 365d/y}\]

5.3.5.2 Operating and Maintenance Cost for Gas Pretreatment System

The GPSs have derived unit O&M costs mostly based on the volume of biogas treated (Table 5-11). The exception is the use of iron salts for H2S removal, in which the unit cost is expressed either $/daily flow of raw wastewater or as $/tonne of VS charged to the digester. The gas pretreatment units in use were specified earlier in the model. Based on the unit costs, the daily and hence annual gas pretreatment costs are calculated for the units in use.
### Table 5-11. Default Values for Unit O&M Costs for Gas Pretreatment Systems.

<table>
<thead>
<tr>
<th>Gas Pretreatment System</th>
<th>Unit O&amp;M Cost (raw biogas unless otherwise specified)</th>
<th>$/scf¹</th>
<th>$/m³ ¹</th>
</tr>
</thead>
<tbody>
<tr>
<td>Iron salts to raw wastewater</td>
<td></td>
<td>0.002</td>
<td>0.077</td>
</tr>
<tr>
<td>Iron salts to digester feed sludge</td>
<td></td>
<td>9.480</td>
<td>10.45</td>
</tr>
<tr>
<td>Iron sponge</td>
<td></td>
<td>0.001</td>
<td>0.036</td>
</tr>
<tr>
<td>Sulfatreat</td>
<td></td>
<td>0.002</td>
<td>0.030</td>
</tr>
<tr>
<td>Binax</td>
<td></td>
<td>0.001</td>
<td>0.033</td>
</tr>
<tr>
<td>Apollo</td>
<td></td>
<td>0.010</td>
<td>0.347</td>
</tr>
<tr>
<td>Activated Carbon</td>
<td></td>
<td>0.0001</td>
<td>0.004</td>
</tr>
</tbody>
</table>

¹Unless otherwise stated

### Annual O&M Cost of GPS

The following equation is valid for only iron salt addition:

\[
AOM_{GPS} = UOM_{GPS} \times DT_S \times Z_{time} \tag{5-62}
\]

Where:
- \( AOM_{GPS} \) = Annual O&M cost of GPS, $/y
- \( UOM_{GPS} \) = Unit O&M cost, $/ton VS ($/tonne VS)
- \( DT_S \) = Daily mass of VS loaded to digester, ton VS/d (tonne VS/d)
- \( Z_{time} \) = Conversion factor for time, 365d/y

Otherwise,

\[
AOM_{GPS} = UOM_{GPS} \times Q_{RB} \times Z_{time} \tag{5-63}
\]

Where:
- \( AOM_{GPS} \) = Annual O&M cost of GPS, $/y
- \( UOM_{GPS} \) = Unit O&M cost of GPS, $/scf RB ($/m³ RB)
- \( Q_{RB} \) = Average daily raw biogas produced, scf RB/d (m³ RB/d)
- \( Z_{time} \) = Conversion factor for time, 365d/y

### 5.3.5.3 Operating and Maintenance Costs for Energy Recovery Systems

#### Unit O&M Cost of ERS

Unit O&M costs for the ERSs are provided in Table 5-12. The unit O&M cost for boilers is based on thermal energy recovered, while the unit O&M costs for direct drive units are based on mechanical energy produced. For the other systems, the unit O&M cost is based on the electrical energy recovered, and are mean values of data recovered from the surveys questionnaires. When any energy recovery system is used, a daily O&M cost is generated, from which the Annual O&M costs for ERSs are derived.
Table 5-12. Unit O&M Costs for Energy Recovery Systems.

<table>
<thead>
<tr>
<th>Technology</th>
<th>Default Unit O&amp;M Costs ($/kWh unless otherwise stated)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler</td>
<td>(5.56x10⁻⁷ $/Btu) 0.0019</td>
</tr>
<tr>
<td>Gas Engine Cogenerator</td>
<td>0.023</td>
</tr>
<tr>
<td>Steam Turbine</td>
<td>0.052</td>
</tr>
<tr>
<td>Gas Turbine</td>
<td>0.033</td>
</tr>
<tr>
<td>Combined Cycle</td>
<td>0.016</td>
</tr>
<tr>
<td>Direct Drive</td>
<td>(1.44$/hp-d) 0.081</td>
</tr>
<tr>
<td>Microturbine</td>
<td>0.020</td>
</tr>
<tr>
<td>Fuel Cell</td>
<td>Molten Carbon DFC300 0.0411</td>
</tr>
<tr>
<td></td>
<td>Phosphoric Acid PC25 0.033</td>
</tr>
<tr>
<td>Stirling Engine</td>
<td>0.008</td>
</tr>
</tbody>
</table>

1. This value is estimated at the same ratio of unit O&M costs to unit capital costs as the PC25

Annual O&M Cost of ERS

The following equation is valid only for boilers

\[
AOM_{ERS} = UOM_{ERS} \times TER \times Z_{energy} \times Z_{time} \quad (5-64)
\]

Where:

- \( AOM_{ERS} \) = Annual O&M cost of boiler, $/y
- \( UOM_{ERS} \) = Unit O&M cost of boiler, $/Btu ($/kWh)
- \( TER \) = Thermal energy recovered, Btu/d (MJ/d)
- \( Z_{energy} \) = Conversion factor for energy, (0.2778kWh/MJ)
- \( Z_{time} \) = Conversion factor for time, 365d/y

The following equation is valid only for direct drive units

\[
AOM_{ERS} = UOM_{ERS} \times MER \times Z_{energy} \times Z_{time} \quad (5-65)
\]

Where:

- \( AOM_{ERS} \) = Annual O&M cost of direct drive unit, $/y
- \( UOM_{ERS} \) = Unit O&M cost of direct drive unit, $/hp-d ($/kWh)
- \( MER \) = Mechanical energy recovered, hp (MJ/d)
- \( Z_{energy} \) = Conversion factor for energy, (0.2778kWh/MJ)
- \( Z_{time} \) = Conversion factor for time, 365d/y

The following equation is valid for the rest of the ERSs
\[ \text{AOM}_{\text{ERS}} = \text{UOM}_{\text{ERS}} \times \text{NEER} \times Z_{\text{time}} \]  \hspace{1cm} (5-66)

Where:
- \( \text{AOM}_{\text{ERS}} \) = Annual O&M cost of ERS, $/y
- \( \text{UOM}_{\text{ERS}} \) = Unit O&M cost of ERS, $/kWh
- \( \text{NEER} \) = Electrical energy recovered, kWh/d
- \( Z_{\text{time}} \) = Conversion factor for time, 365d/y

5.3.5.4 Total Annual O&M Cost

The total annual O&M costs is calculated as the sum of the annual O&M costs for the anaerobic digestion system, the GPS, and the ERS.

\[ \text{AOMC} = \text{AOM}_{\text{AD}} + \text{AOM}_{\text{GPS}} + \text{AOM}_{\text{ERS}} \]  \hspace{1cm} (5-67)

Where:
- \( \text{AOMC} \) = Total annual O&M cost, $/y
- \( \text{AOM}_{\text{AD}} \) = Annual O&M cost of digester, $/y
- \( \text{AOM}_{\text{GPS}} \) = Annual O&M cost of GPS, $/y
- \( \text{AOM}_{\text{ERS}} \) = Annual O&M cost of ERS, $/y

5.3.6 Annual Energy Costs

The annual energy costs result from expenditures for purchased electricity generated offsite, and for fuels including natural gas and diesel. For electricity costs, the model user has previously specified the purchased electricity and prices for regular and peak usage. To complete the cost estimate, the model user then must enter the percent of time that electricity is consumed at regular pricing.

With respect to natural gas and diesel fuel, the annual costs are calculated as the product of the annual natural gas and diesel demands of the plant, and the prices paid by the plant for the natural gas and diesel. For natural gas in particular, a tiered pricing is allowed to account for a pricing structure that may change after a certain threshold volume consumed is reached. The model user must define the threshold value.

The annual energy cost is the sum of the annual costs for electricity, natural gas and diesel fuel.

5.3.6.1 Annual Electricity Cost

\[ \text{AelC} = [\%_{\text{dem-normal}} \times P_{\text{elec-normal}} + \%_{\text{dem-peak}} \times P_{\text{elec-peak}} ] \times \text{AelD} \]  \hspace{1cm} (5-68)

Where:
- \( \text{AelC} \) = Annual electricity cost, $/y

5-34
AelD = Annual electricity demand, kWh/y

\[ P_{\text{elec-normal}} = \text{Normal price of electricity, } \$/\text{kWh} \]

A default value of 0.09$/kWh was assumed.

\[ P_{\text{elec-peak}} = \text{Peak price of electricity, } \$/\text{kWh} \]

Value is supplied by model user (e.g. 0.15 $/kWh).

\[ \%_{\text{dem-normal}} = \text{Percentage of annual electricity demand for normal electricity price, } \% \]

Value is supplied by model user (e.g. 80%).

\[ \%_{\text{dem-peak}} = \text{Percentage of annual electricity demand for peak electricity price, } \% \]

Value is calculated based on user input for percent normal electricity price.

### 5.3.6.2 Annual Natural Gas Cost

\[ \text{AngC} = P_{\text{ngas}} \times \text{AngD} \quad (5-69) \]

Where:

\[ \text{AngC} = \text{Annual natural gas cost, } \$/\text{y} \]

\[ \text{AngD} = \text{Annual natural gas demand, scf/y (m}^3/\text{y}) \]

\[ P_{\text{ngas}} = \text{Price of natural gas, } \$/\text{scf (}/$\text{m}^3) \]

A default value of 0.007 $/scf (0.25 $/m³) was assumed.

### 5.3.6.3 Annual Diesel Cost

\[ \text{AdiC (}/$/\text{year}) = P_{\text{diesel}} \times \text{AdiD} \quad (5-70) \]

Where:

\[ \text{AdiC} = \text{Annual diesel cost, } \$/\text{y} \]

\[ \text{AdiD} = \text{Annual diesel demand, gal/y (L/y)} \]

\[ P_{\text{diesel}} = \text{Price of diesel paid by plant, } \$/\text{gal (}/$/\text{L}) \]

A default value of 1.5 $/gal (0.8 $/L) was assumed.

### 5.3.6.4 Annual Energy Cost

\[ \text{AEC} = \text{AelC} + \text{AngC} + \text{AdiC} \quad (5-71) \]

Where:

\[ \text{AEC} = \text{Annual energy cost total, } \$/\text{y} \]

\[ \text{AelC} = \text{Annual electricity cost, } \$/\text{y} \]

\[ \text{AngC} = \text{Annual natural gas cost, } \$/\text{y} \]

\[ \text{AdiC} = \text{Annual diesel cost, } \$/\text{y} \]
5.3.7 **Net Annual Operating Cost**

The net annual operating cost is the sum of the annual O&M costs and the annual energy costs. This is shown in Equation 5-72.

\[
\text{AOC} = \text{OMC} + \text{AEC} \quad (5-72)
\]

Where:

\[
\begin{align*}
\text{AOC} & = \text{Net annual operating cost total, } \$/y \\
\text{AOMC} & = \text{Annual O&M cost, } \$/y \\
\text{AEC} & = \text{Annual energy cost, } \$/y
\end{align*}
\]

5.3.8 **Economic Assessment Factors**

A number of economic assessment factors are determined to assist the model user in evaluating the cost-effectiveness of the various combinations of anaerobic digestion, GPSs and ERSs. These assessment parameters include:

- Payback period
- Net savings
- Net total revenue
- Net operating cost
- Net present value
- Net total annualized economic cost and internal rate of return

These economic parameters will be discussed individually.

5.3.8.1 **Payback Period**

Two procedures are available for calculating the payback period. One is a simple payback period, and is the initial capital cost divided by the quantity of combined annual savings and revenues less the annual operating costs. If the annual costs exceed the revenues and savings a negative payback is calculated.

*Simple Payback*

\[
PBP = \frac{\text{CC}}{\text{AR} + \text{AS} - \text{AOC}} \quad (5-73)
\]

*Simple Payback with Replacement Capital*

\[
PBP = \frac{\text{CC} + \text{RC}}{\text{AR} + \text{AS} - \text{AOC}} \quad (5-74)
\]
Where:

- **PBP** = Payback period, y
- **AR** = Net annual revenue, $/y
- **AS** = Net annual saving, $/y
- **AOC** = Net annual operating cost, $/y
- **CC** = Initial capital cost, $
- **RC** = Replacement cost, $
- **i** = Discount rate, %

A default value of $i = 6\%$ was assumed.

The payback period with replacement cost is more reflective of the benefit when the life cycle costing approach is used, whereas the simple payback period is not.

### 5.3.8.2 Net Saving

The Net Saving is the value of the annual savings accrued over the specified planning period, adjusted for the discount rate of money applied over the period.

\[
NS = AS \frac{(1 + i)^N - 1}{i \times (1 + i)^N} 
\]

(5-75)

Where:

- **NS** = Net saving, $
- **AS** = Net annual saving, $/y
- **N** = Planning period, y

A default value of $N = 20$ years was assumed.

### 5.3.8.3 Net Total Revenue

The Net Total Revenue is the value of the total revenues obtained from the sales of energy (heat, electricity and biogas) to offsite customers accrued over the specified planning period, adjusted for the discount rate of money applied over the period.

\[
NTR = ATR \frac{(1 + i)^N - 1}{i \times (1 + i)^N} 
\]

(5-76)

Where:

- **NTR** = Net total revenue, $
- **ATR** = Net annual total revenue, $/y
N = Planning period, y
A default value of N = 20 years was assumed.
i = Discount rate, %
A default value of i = 6% was assumed.

5.3.8.4 Net Operating Cost
The Net Operating Cost is the value of the annual operating costs incurred over the specified planning period, which is adjusted for the discount rate of money applied over the period.

\[ \text{NOC} = \text{AOC} \times \frac{(1 + i)^N - 1}{i \times (1 + i)^N} \]  

(5-77)

Where:
NOC = Net operating cost, $
AOC = Net annual operating cost, $/y
N = Planning period, year.
A default value of N = 20 years was assumed.
i = Discount rate, %
A default value of i = 6% was assumed.

5.3.8.5 Net Present Value (Net Life Cycle Cost)
The Net Present Value, or Net Life Cycle Cost is the difference in value after net savings and net total revenues have been deducted from the net capital and net operating costs.

\[ \text{LCC} = \text{CC} + \text{NOC} - \text{NS} - \text{NTR} \]  

(5-78)

Where:
LCC = Net life cycle costs or net present value, $
CC = Net capital cost, $
NOC = Net operating cost, $
NS = Net saving, $
NTR = Net total revenue, $

5.3.8.6 Net total annualized economic cost and internal rate of return

Net Annualized Capital Cost
The Net Annualized Capital Cost expresses the net capital cost on a constant annual rate accounting for the planning period and the discount rate of money.

\[ \text{ACC} = \frac{\text{CC} \times \frac{i \times (1 + i)^N}{(1 + i)^N - 1}} \]  

(5-79)
Where:
ACC  = Net annualized capital cost, $/y
CC   = Net capital cost, $
N    = Planning period, y
A default value of N = 20 years was assumed.
i    = Discount rate, %
A default value of i = 6% was assumed.

Net Total Annualized Economic Cost
This economic parameter is the difference in value after net annual savings and net annual total revenues have been deducted from the net annualized capital and net annual operating costs. It corresponds to the Net Life Cycle Cost, except the Net Life Cycle Cost is expressed in terms of constant dollars, while the Net Annualized Economic Cost is expressed as constant dollars per year over the planning period.

TAEC = ACC + AOC - ATR - AS                   (5-80)

Where:
TAEC = Net total annual economic cost, $/y
ACC  = Net annualized capital cost, $/y
AOC  = Net annual operating cost, $/y
AS   = Net annual saving, $/y
ATR  = Net annual total revenue, $/y

Internal Rate of Return
The total life cycle cost discussed above uses the discount rate and the planning period. The internal rate of return is the discount rate that just makes the life cycle cost equal to zero for a given time frame of interest.

The internal rate of return (IRR) can be estimated iteratively with the LCAMER spreadsheet model if the life cycle cost calculated by the model is negative. [If the life cycle cost is positive, the internal rate of return has no practical meaning]. To estimate the IRR in the spreadsheet model, the following steps are required:

♦ Enter a time frame of interest (this must be greater than the simple payback period).
♦ Enter an estimated IRR.
♦ Compare the P/A factor calculated to the target value in the line below.
♦ Increase or decrease the estimated IRR value (%) until the P/A factor calculated is approximately equal to the target value. At that point, the estimated value can be taken as the IRR for the project.

Table 5-13 shows how this is shown in the LCAMER spreadsheet model.

<table>
<thead>
<tr>
<th>Column</th>
<th>Row</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>246</td>
<td>ESTIMATING INTERNAL RATE OF RETURN-IRR- (%) (only valid when LCC &lt; 0)</td>
</tr>
<tr>
<td></td>
<td>247</td>
<td>Select a time frame of interest (y): Must be greater than Simple Payback Period</td>
</tr>
<tr>
<td></td>
<td>248</td>
<td>Select an IRR value (%): Adjust until P/A Factor = Value to Achieve</td>
</tr>
<tr>
<td></td>
<td>249</td>
<td>P/A Factor: Change IRR until this equals Target Value to Achieve</td>
</tr>
<tr>
<td></td>
<td>250</td>
<td>Target Value to Achieve</td>
</tr>
<tr>
<td></td>
<td>251</td>
<td>3.21</td>
</tr>
<tr>
<td></td>
<td>252</td>
<td>1.97</td>
</tr>
</tbody>
</table>

Note that the IRR calculated here is only an estimate. Because of limitations in the spreadsheet model set-up, the IRR includes only initial capital, not replacement costs. Additionally, IRR can be a difficult parameter to appropriately compare between projects, depending on the size of the capital expenditures and time scales involved. For this spreadsheet model, we believed that the initial capital payback period, the total life cycle cost and the net total annualized economic cost are better comparison parameters than the IRR.

5.4 Listing of Parameters

5.4.1 Listing of Units of Measure

U.S. Based Units (used in U.S. spreadsheet model)

- $ = US Dollars
- % = Percentage
- %VS = Percent volatile solids
- °F = Degrees Fahrenheit
- Btu = British thermal units
- cap = Capita (person)
- d = Days
- ft = Feet
- gpd = Gallons per day
- h = Hour
- hp = Horsepower
- L = Litre
- MGD = Million gallons per day
- scf = Standard cubic feet
- ton = Ton
- y = Year

Metric Based Units (used in metric spreadsheet model)

- $ = US Dollars
5.4.2 Listing of Parameters used in Technical Functions

\(\alpha\) = Proportion coefficient for primary sludge production (dimensionless)

\(\rho_{BP}\) = Density of biological sludge, lb/gal (kg/m\(^3\))

\(\rho_{SP}\) = Density of primary sludge, lb/gal (kg/m\(^3\))

\(\%CH_4_{PB}\) = Percentage of CH\(_4\) in purified biogas, %

\(\%CH_4_{RB}\) = Percentage of CH\(_4\) in raw biogas, %

\(\%CO_2_{PB}\) = Percentage of CO\(_2\) in purified biogas, %

\(\%CO_2_{RB}\) = Percentage of CO\(_2\) in raw biogas, %

\(\%TS_{th}\) = Total solids concentration in the thickened sludge, %

\(\%VSS\) = Fraction of mixed liquor suspended solids that are volatile (decimal fraction)

\(\%VS_{th}\) = Percent volatile solids in the thickened sludge, %

\(A\) = Average surface of digester, ft\(^2\) (m\(^2\))

\(AFPP\) = Average flow per person, gal/cap-d (m\(^3\)/cap-d)

\(c\) = Average specific heat of sludge, Btu/lb/oF (J/kg/oC)

\(C_{ERS}\) = Capacity of energy recovery of energy recovery system, kW

\(C_{SB}\) = Concentration of biological sludge entering the digester, %VS

\(C_{SP}\) = Concentration of primary sludge entering the digester, %VS

\(D_{CO_2}\) = Density of CO\(_2\) in digester gas, lb/scf (kg/m\(^3\))

\(D_{CH_4}\) = Density of CH\(_4\) in digester gas, lb/scf (kg/m\(^3\))

\(D_{SP}\) = Density of primary sludge in wastewater, lb/gal (kg/m\(^3\))

\(D_{TS}\) = Daily total tonne VS treated, ton/d (tonne/d)
DTBS = Daily tonne VS of biological sludge, ton/d (tonne/d)
DTPS = Daily tonne VS of primary sludge, ton/d (tonne/d)
eI = Electricity purchased by a WWTP, kWh/y
eDM = Unit energy needed for digester mixing, W/gal (W/m³)
ECH4 = Emission of methane from the boiler, tonCH4/y (tonneCH4/y)
ECO2 = Emission of carbon dioxide from the boiler, tonCO2/y (tonneCO2/y)
EDH = Energy needed for digester heating, Btu/h (W)
EDM = Energy needed for digester mixing, W
E offset CO = Offset emission of CO, tonCO/y (tonneCO/y)
E offset CO2 = Offset emission of CO2, tonCO2/y (tonneCO2/y)
E offset NOX = Offset emission of NOX, tonNOx/y (tonneNOX/y)
E offset PM10 = Offset emission of PM10, tonPM10/y (tonnePM10/y)
E offset SO2 = Offset emission of SO2, tonSO2/y (tonneSO2/y)
E offset VOC = Offset emission of VOC, tonVOC/y (tonneVOC/y)
ESO2 = Emission of SO2 from boiler, tonSO2/y (tonneSO2/y)
ECboiler = Combustion efficiency of the boiler (as decimal fraction)
ECCH4 = Energy content of methane, default 961.5 Btu/scf (35.8MJ/m³)
ECngas = Energy content of natural gas, default 1030Btu/scf (38.3MJ/m³)
ECE = Energy recovery efficiency, %
ECEMech = Mechanical energy recovery efficiency, %
ECETher = Thermal energy recovery efficiency, %
epkCO,ET J = Amount of CO produced per electricity type in a Jurisdiction, tonCO/kWh (tonneCO/kWh)
epkCO2,ET J = Amount of CO2 produced per electricity type in a Jurisdiction, tonCO2/kWh (tonneCO2/kWh)
EpkNOX,ET J = Amount of NOX produced per electricity type in a Jurisdiction, tonNOx/kWh (tonneNOX/kWh)
epkPM10,ET J = Amount of PM10 produced per electricity type in a Jurisdiction, tonPM10/kWh (tonnePM10/kWh)
epkSO2,ET J = Amount of SO2 produced per electricity type in a Jurisdiction, tonSO2/kWh (tonneSO2/kWh)
epkVOC,ET J = Amount of VOC produced per electricity type in a Jurisdiction, tonVOC/kWh (tonneVOC/kWh)
FET J = Fraction of electricity type produced in a given Jurisdiction, %
$F_{v,j}$ = Conversion factor from VSS to COD units, lb$_{COD}$/lb$_{VSS}$ (kg$_{COD}$/kg$_{VSS}$)

GPR = Gas production rate, scf/lb$_{VSadded}$ (m$^3$/kg$_{VSadded}$)

HEE = Heat exchanger efficiency, %

$K_{1,T,c,j}$ = First order substrate consumption for sludge j at temperature T and process condition c, d$^{-1}$

$K_{h,T,c,j}$ = First order hydrolysis constant for sludge j at temperature T and process condition c, d$^{-1}$

j = Sludge type: j = 1 for primary sludge; j = 2 for WAS

M$_{SO_{2}}$ = Molecular weight of SO$_{2}$, lb/mol (g/mol)

MER = Mechanical energy recovered, hp (MJ/d)

MLSS = the concentration of mixed liquor suspended solids, mg/L (kg/m$^3$)

N = Numbers of digesters

NTER = Net thermal energy recovered, Btu/d (MJ/d)

OLR = Organic loading rate, ton$_{VS}$/gal-d (tonne$_{VS}$/m$^3$-d)

P = Population served, number of people

ppm H$_2$SPB = H$_2$S in purified biogas, ppm

ppm H$_2$SRB = H$_2$S in raw biogas, ppm.

Q$_{DRBEE}$ = Total daily raw biogas for energy production, scf/d (m$^3$/d)

Q$_{PB}$ = Daily purified biogas produced, scf/d (m$^3$/d)

Q$_{PB}$E = Amount of daily total purified biogas for energy production, scf/d (m$^3$/d)

Q$_{PB}$EE = Amount of daily purified biogas for electrical energy production, scf/d (m$^3$/d)

Q$_{PB}$ME = Amount of daily purified biogas for mechanical energy production, scf/d (m$^3$/d)

Q$_{PB}$N = Amount of daily purified biogas not used, scf/d (m$^3$/d)

Q$_{RB}$ = Daily raw biogas produced, scf/d (m$^3$/d)

Q$_{RB}$EE = Daily raw biogas for electrical energy production, scf/d (m$^3$/d)

Q$_{RB}$TE = Daily raw biogas for thermal energy production, scf/d (m$^3$/d)

Q$_{S}$ = Daily total sludge flow to digesters, scf/d (m$^3$/d)

Q$_{SB}$ = Biological sludge flow, scf/d (m$^3$/d)

Q$_{SP}$ = Primary sludge flow, scf/d (m$^3$/d)

Q$_{W}$ = Daily wastewater flow, scf/d (m$^3$/d)

Q$_{wsh}$ = The thickened sludge flow, scf/d (m$^3$/d)

RCO$_2$ = CO$_2$ Pretreatment Efficiency, %

RH$_2$S = H$_2$S Pretreatment Efficiency, %
\[ \text{R}_{\text{W/T}} = \text{Ratio working volume total volume, scf/scf (m}^3/\text{m}^3) \]
\[ \text{R}_{\text{G/W}} = \text{Ratio of grit to raw wastewater in primary sludge, gal/gal (m}^3/\text{m}^3) \]
\[ S_{H,i-1,j} = \text{Soluble concentration of the constituents entering stage } i, \text{ mgCOD/L} \] 
\[ \left(\text{kg COD/m}^3\right) \]
\[ \text{SRT}_{\text{aerobic}} = \text{Solids retention time of the biological treatment system, days} \]
\[ \text{SRT}_i = \text{Design sludge retention time of stage } i, \text{ d} \]
\[ \text{SRT}^*_i = \text{Hypothetical sludge retention time of stage } i, \text{ d} \]
\[ T_1 = \text{Temperature of sludge to be treated, } ^\circ\text{F (}^\circ\text{C)} \]
\[ T_2 = \text{Temperature of sludge from digester, } ^\circ\text{F (}^\circ\text{C)} \]
\[ T_3 = \text{Average ambient temperature per jurisdiction, } ^\circ\text{F (}^\circ\text{C)} \]
\[ \text{TER} = \text{Thermal energy recovered, Btu/d (MJ/d)} \]
\[ U = \text{Heat transfer coefficient, Btu/h-ft}^2-^\circ\text{F (W/m}^2-^\circ\text{C)} \]
\[ V = \text{Aeration basin volume, gal (m}^3) \]
\[ V_T = \text{Total volume of digesters, gal (m}^3) \]
\[ V_{\text{chemical}} = \text{Volume of chemical added per day, scf/d (m}^3/\text{d)} \]
\[ V_W = \text{Working volume of digesters, gal (m}^3) \]
\[ V_{\text{SL}i} = \text{Load of VS leaving stage } i, \text{ lbVS/d (kgVS/d)} \]
\[ V_{\text{SL}0} = \text{Load of VS entering stage 1, lbVS/d (kgVS/d)} \]
\[ \text{VSR} = \text{Volatile solids reduction, } \% \]
\[ V_{\text{SS}i-1,j} = \text{Volatile suspended solid concentration of sludge } j \text{ entering stage } i, \text{ mg/L (kg/m}^3) \]
\[ V_{\text{SS}wasted} = \text{Daily mass of biological volatile solids wasted, lb/day (kg/day)} \]
\[ x = \text{Variable} \]
\[ Z = \text{Conversion factor} \]

5.4.3 Listing of Parameters used in Economic Functions

\[ \%_{\text{dem-normal}} = \text{Percentage of annual electricity demand for normal electricity price, } \% \]
\[ \%_{\text{dem-peak}} = \text{Percentage of annual electricity demand for peak electricity price, } \% \]
\[ \%_{\text{pro-normal}} = \text{Percentage of annual produced electricity for normal electricity price, } \% \]
\[ \%_{\text{pro-peak}} = \text{Percentage of annual produced electricity for peak electricity price, } \% \]

\[ \text{AdiC} = \text{Annual diesel cost, } $/\text{y} \]
\[ \text{AdiD} = \text{Annual diesel demand, gal/y (L/y)} \]
\[ \text{ACC} = \text{Net annualized capital cost, } $/\text{y} \]
AEC  = Annual energy cost total, $/y
AelC  = Annual electricity cost, $/y
AelP  = Annual offsite electricity purchased (demand for offsite electricity), kWh/y
AngC  = Annual natural gas cost, $/y
AngD  = Annual natural gas demand, scf/y (m³/y)
AOC   = Net annual operating cost total, $/y
AOMAD = Annual operating and maintenance cost of digester, $/y
AOMERS= Annual operating and maintenance cost of ERS, $/y
AOMGPS= Annual operating and maintenance cost of GPS, $/y
AOMC  = Annual O&M cost, $/y
AR    = Net annual revenue, $/y
AS    = Annual net saving, $/y
ASVS  = Annualized salvage value, $/y
ATR   = Net annual total revenue, $/y
crGHG = Greenhouse gas emission credits, $/ton ($/tonne)
CGPS  = Unit cost of gas pretreatment system (iron salts), $/ton VS/d ($/tonne VS/d)
CPERS = Capacity of ERS, kW; hp; Btu/hr
CC    = Net capital cost, $
DTs   = Digester feed solids mass, ton VS/d (tonne VS/d)
EoffsetCO2 = Offset emission of CO2, ton CO2/y (tonne CO2/y)
ECgas = Energy content of natural gas, 1030 Btu/scf (38.3 MJ/m³)
G     = Grant received, $
i    = Discount rate, %
ICAD  = Initial capital cost of anaerobic digester, $
ICERS = Initial capital cost of ERS, $
ICGPS = Initial capital cost of gas pretreatment system, $
LCC   = Net life cycle costs or net present value, $
MER   = Mechanical energy recovered, hp (MJ/d)
N     = Planning period, y
NOC   = Net operating cost, $
NR    = Round number of replacement of GPS during the planning period
NS    = Net saving, $
NTER  = Net thermal energy recovered, Btu/d (MJ/d)
NTR = Net total revenue, $
OPBE = Onsite produced bioelectricity, kWh/d
P_{bioelec} = Selling price of bio-electricity, $/kWh
P_{bioheat} = Selling price of bioheat, $/Btu ($/kJ)
P_{biogas} = Selling price of the biogas, $/scf ($/m^3)
P_{diesel} = Price of diesel paid by plant, $/gal ($/L)
P_{elec-normal} = Normal electricity price paid by the plant, $/kWh
P_{elec-peak} = Peak electricity price paid by the plant, $/kWh
P_{gas} = Price of natural gas paid by the plant, $/gal ($/m^3)
PBP = Payback period, y
Q_{bioelec-s} = Annual produced and plant-consumed electricity, kWh/y
Q_{bioheat-s} = Annual plant-consumed bioheat, Btu/y (kJ/y)
Q_{PB} = Daily raw biogas for pre-treatment, scf/d (m^3/d)
Q_{PBN} = Amount of daily purified biogas not used, scf/d (m^3/d)
Q_{surbioelec} = Surplus BioElectricity, kWh/y
Q_{W} = Daily wastewater flow, MGD (m^3/d)
RCGPS = Replacement cost of GPS, $
RS = Replacement schedule of GPS, y
RS_{ERS} = Replacement schedule for ERS, y
SV_S = Net salvage value total, $
SV_{ERS} = Net salvage value of ERS, $
SV_{LE} = Net salvage value of equipment with service life longer than planning period, $
TAEC = Net total annual economic cost, $/y
TER = Thermal energy recovered, Btu/d (MJ/d)
UC_{AD} = Unit capital cost of anaerobic digester, $/gal ($/m^3)
UC_{ERS} = Unit capital cost of ERS, $/kW; $/hp; $/Btu/h
UC_{GPS} = Unit capital cost of gas pretreatment system (iron salts), $/scf_{rw} \text{, } ($/m^3_{rb}
\text{d})
UC_{MT} = Unit cost of microturbine in 30 – 120 kW capacity range, $/kW
UOM_{AD} = Unit O&M costs of digester, $/ton_{VS} ($/tonne_{VS})
UOM_{ERS} = Unit O&M cost of ERS, $/kWh; $/hp-h; $/Btu
UOM_{GPS} = Unit O&M cost (iron salts), $/ton_{VS} ($/tonne_{VS})
\text{= Unit O&M cost of GPS, $/scf}_{RB} ($/m^3_{RB})
$V_T = \text{Total volume of digester, gal (m}^3\text{)}$

$Z = \text{Conversion factor}$
CHAPTER 6.0

LCAMER USER MANUAL AND CASE STUDIES

6.1 Overview of Spreadsheet for Digester Energy Evaluation

This manual is intended to provide you with a step-by-step guide to working through the spreadsheet. When necessary, specific points of theory or equations will be mentioned, but this is kept to a minimum. For detailed discussion of the theory behind the calculations being conducted, please see the project report.

There are seven worksheets in the spreadsheet. The first worksheet, Info to Users, briefly describes the contents of each worksheet. The Technical and Economics worksheets are the ones that you will use to conduct your analysis. They will be described in more detail shortly.

The remaining four worksheets contain calculations and data that are used by the Technical and Economics worksheets. You may look at these spreadsheets, but do not enter anything into them. As will be described later, if you would like to input your own estimates and not use the information contained in these supporting worksheets, you will do that directly in the Technical and Economics worksheets.

Briefly, the four supporting worksheets contain the following information. For more information about the details of the calculations, please refer to the project report. “AD-Technology” worksheet calculates volatiles solids reduction estimates for the types of anaerobic digestion technologies considered. “Default Temp” worksheet contains annual average ambient temperatures for all 50 states, the District of Columbia, and the 10 Canadian provinces. Also included in this worksheet are the average emissions due to electricity generation in each jurisdiction. “ERS Emission Factors” worksheet contains the emissions factors for estimating the air emissions from using energy recovery on-site at your wastewater treatment plant. Finally, “Factor-ERS” worksheet contains intermediate calculations for default unit capital costs as well as factors for replacement capital costs.

6.2 Before You Begin – A Precautionary Note

If you follow the guidelines below, you should not encounter any problems with the spreadsheet. The manual attempts to clearly indicate which cells require your entry of information and which cells should not be touched. Nevertheless, the possibility remains that you
may inadvertently enter information into a cell that previously contained an equation (this we
know from painful experience). **THEREFORE, WE HIGHLY RECOMMEND THAT YOU
KEEP THE MASTER SPREADSHEET UNTouched AND ALWAYS PREPARE A
COPY WHEN YOU CONDUCT YOUR EVALUATIONS.** This way, if a mistake occurs,
you can simply prepare another copy of the master spreadsheet. Additionally, you may want to
prepare separate copies for every condition you examine because the spreadsheet only produces
one set of results for every input set.

### 6.3 Technical Worksheet

#### 6.3.1 Using the Technical Worksheet

The starting place for using the spreadsheet is the Technical worksheet. What follows is a
row-by-row description of the worksheet, indicating where you should input information, where
to find results, and where to not modify anything.

**6.3.1.1 Rows 1-8**

The first 8 rows provide the color-coded legend. Of the 7 colors shown, the one that
matters most to you is light-green (this may also look light-blue). Cells that are labeled this color
require input from you. The rest of the colors were more useful to us when we prepared the
spreadsheet, but they also serve to help break up the monotony of row upon row of numbers.

**6.3.1.2 Rows 10-13**

*Do not modify these rows.* They show the results of calculations later in the spreadsheet
which must be used to calculate the VS reduction (%VSR). If you do not like the numbers
showing in these rows, be patient. The chance to make changes occurs later.

#### 6.3.2 The Anaerobic Digestion Technology and VSR

**6.3.2.1 Rows 16-18: Selecting Anaerobic Digestion Technology**

The spreadsheet considers 8 different anaerobic digestion technologies described in Table
7-1. Select one and only one of these technologies by typing 'yes' into the appropriate cell of
Row 16 (B16-I16), and 'no' into the other 7 cells. If you type in more than one yes, the default
calculations throughout will be wrong.

Next, in Row 17, in the column under the selected digestion technology, type in the SRT
in days for the first stage. If you are considering a 2-stage system, type in the second stage SRT
in Row 18. The only column that you need worry about is the one you selected in Row 16. A 'no'
entered into a cell in Row 16 turns off the calculations associated with all other cells. For
example, if you selected 'yes' for Meso-2-Stage (Cell C16), then you only need to enter SRT
values in Cells C17 and C18. None of the other cells in Rows 17 and 18 will be used.
If you are interested in evaluating a different anaerobic digestion technology than those noted, select either Meso-1-Stage if you are considering a 1-stage system or Meso-2-Stage if you are considering a 2-stage system. Then enter the appropriate SRTs for the stages in Rows 17 and 18. You will then enter your own VS Reduction (%VSR) later.

### Table 6-1. Anaerobic Digester Options.

<table>
<thead>
<tr>
<th>Abbreviation in the Spreadsheet</th>
<th>Technology</th>
</tr>
</thead>
<tbody>
<tr>
<td>Meso-1-Stage</td>
<td>One stage, mixed mesophilic (35-37°C) digester</td>
</tr>
<tr>
<td>Meso-2-Stage</td>
<td>Two sequential stages, each stage is a mixed mesophilic digester</td>
</tr>
<tr>
<td>Meso-Meso Acid Phase</td>
<td>Two sequential stages, first stage is a mixed, mesophilic, acid-phase digester; second stage is a mixed, mesophilic digester</td>
</tr>
<tr>
<td>Meso-Thermo Acid Phase</td>
<td>Two sequential stages, first stage is a mixed, mesophilic, acid-phase digester; second stage is a mixed, thermophilic digester (see Thermo-1-Stage)</td>
</tr>
<tr>
<td>Thermo-1-Stage</td>
<td>One stage, mixed thermophilic (55-60°C) digester</td>
</tr>
<tr>
<td>Thermo-2-Stage</td>
<td>Two sequential stages, each stage is a mixed thermophilic digester</td>
</tr>
<tr>
<td>TPAD (Meso-Thermo)</td>
<td>Two sequential stages, first stage is a mixed, mesophilic, digester; second stage is a mixed, thermophilic digester</td>
</tr>
<tr>
<td>TPAD (Thermo-Meso)</td>
<td>Two sequential stages, first stage is a mixed, thermophilic, digester; second stage is a mixed, mesophilic digester</td>
</tr>
</tbody>
</table>

#### 6.3.2.2 Rows 19-20

**Do not modify these rows.** Row 19 calculates the total SRT for your selected system and Row 20 calculates the VS Reduction (%VSR) that the spreadsheet is calculating in the AD-Technology worksheet. If you don't like the %VSR number shown, don't worry, you can override it later (see Rows 22-23 next).

#### 6.3.2.3 Rows 22-23: Selecting VSR

In Cell B22 enter 'yes' if you would like the spreadsheet to use the default %VSR that is calculated in the AD-Technology sheet and shown in Row 20. Select 'no' if you would like to enter your own estimate in Cell B23. If you enter 'yes' in Cell B22, you may ignore Cell B23.
**Suggestion:** Unless you have a high level of confidence with your VSR estimate, we recommend using our default calculation. This is based on an extensive review of full-scale systems.

**Note:** The %VSR shown in Row 20 will not accurately reflect our estimated %VSR until you complete the information requested through Row 100. So do not be alarmed the first time through the spreadsheet if the VSR does not seem accurate. Once you have completed the requested information, come back to Row 20 and you will see the correct value. You may nevertheless still wish to override it with Cell B22. Sorry about the iteration.

6.3.2.4 **Rows 24-28**

*Do not modify these rows.* The information in these rows is used later. Cell B25 shows the actual %VSR that will be used (either the default calculated value or your entered value). Cell B28 shows the expected biogas yield in m³/kg VS added to the digesters. These values will not be completely accurate until information is entered through Row 100.

**6.3.3 Wastewater Flow**

**6.3.3.1 Rows 31-32**

If you wish to use our default calculation for wastewater flow, enter 'yes' in Cell B31. To enter your own wastewater flow, enter 'no' in Cell B31 and then enter your wastewater flow into Cell B32.

**Suggestion:** We highly recommend that you enter 'no' in Cell B31 and enter your wastewater flow into Cell B32. Our default calculation is based on standard per capita wastewater generation numbers.

**Note:** Be sure to enter your flow as cubic meters per day (m³/day). Conversion factors are provided elsewhere to help you make the conversion.

**6.3.3.2 Rows 35-36**

Enter your population serviced in Cell B35 and the type of sewer system, either 'separate' or 'combined' in Cell B36 if you want the default calculation conducted. If you have entered your own wastewater flow, you may ignore these cells.

**6.3.3.3 Rows 37-41**

*Do not modify these rows.* Cell B41 shows the wastewater flow that will be used by the spreadsheet based on the input provided in Rows 31-36.
6.3.4 Primary Sludge Flows and Solids Concentrations

6.3.4.1 Rows 44-45: Selecting Primary Sludge Flows

If you wish to use our default calculation for primary flow, enter 'yes' in Cell B44. To enter your own primary sludge flow, enter 'no' in Cell B44 and then enter your wastewater flow into Cell B45. Be sure to use cubic meters per day.

**Suggestion:** We highly recommend that you enter 'no' in Cell B44 and enter your primary sludge flow into Cell B44. The results of the spreadsheet are sensitive to sludge flow rates so these numbers should be as accurate as possible.

6.3.4.2 Rows 48-51

If you have entered your own primary sludge flow, ignore these rows. If you would like the default primary sludge flow, choose one of the primary treatment choices from Row 49 [none, natural (setting without chemical addition), or chemical] and type that into Cell B48. Then in Row 50, enter the percent of total wastewater flow entering the primary clarifier that will leave with the primary sludge. We recommend 0% for none, 1% for natural and 3% for chemical. If and only if you have selected 'chemical' in Cell B48, then enter the chemical dosage volume (cubic meters per day) into Cell D51. Otherwise you may ignore Row 51.

6.3.4.3 Rows 53-58

**Do not modify these rows.** Cell B58 shows the primary sludge flow that will be used by the spreadsheet based on the input provided in Rows 44-51.

6.3.4.4 Rows 60-69: Thickening Primary Sludge

**Do not modify Rows 65 or 69.** If you do not thicken primary sludge, follow these instructions. Enter 'no' into Cell B60 and then enter the value shown in Cell B58 into Cell B61. Skip Rows 63-65 and then enter the Percent Total Solids (%TS) in Cell B67 and the percent of total solids that are volatile solids into Cell B68. Whether you chose the default in Cell B44 or not, these instructions will work.

If you do thicken primary sludge, and you know the flow rate, enter 'no' into Cell B60, enter your flow rate (cubic meters per day) into Cell B61 and enter the appropriate values into Cells B67 and B68 as described in the previous paragraph.

If you thicken primary sludge and wish to depend on our default calculations, enter 'yes' into Cell B60, ignore Cell B61 and then enter information for Cells B63 and B64 (the appropriate solids concentrations before thickening) and also in Cells B67 and B68 (the appropriate solids concentrations after thickening).
6.3.4.5 Rows 71-72

**Do not modify these rows.** Cell B71 shows the thickened primary sludge flow that will be used by the spreadsheet and Cell B72 shows the mass flow rate of VS that is due to the primary sludge.

6.3.5 Biological Sludge Flows and Solids Concentrations

6.3.5.1 Rows 76-86: Estimating Biological Sludge Flows

**Do not modify Rows 82, 84 or 86.** Enter the solids retention time (SRT, in days) of your biological treatment process into Cell B76. Then enter your aeration volume (cubic meters) into Cell B77. In Cell B78, we calculate the hydraulic retention time (HRT, in hours) based on Cells B76 and B77. If this number doesn't agree with your HRT, then double-check your entries in B76 and B77.

In Cell B79, enter your mixed liquor suspended solids concentration (MLSS, mg/L) and in Cell B80, enter the percentage of your MLSS that is volatile. In Cell B81, we estimate your system's MLSS using the information in Cells C81, D81 and E81. This estimate is just to provide a comparison. If the value you entered in Cell B79 is very different from our estimate, you may wish to double-check your entries. Alternatively, you may wish to use our number in Cell B81 as a default MLSS. To do this, type the value shown in Cell B81 into Cell B79 (our number in Cell B81 is not used in subsequent calculations so if you want to use the number, you must enter into Cell B79.)

Enter your *thickened* biological sludge total solids concentration (%TS) into Cell B83.

6.3.5.2 Rows 88-92: Selecting Biological Sludge Flows

**Do not modify Rows 90 or 92.** If you would like to use the calculated default thickened biological sludge flow in Cell B86, enter 'yes' in Cell B88 and ignore Cells B89 and B91. If you know your thickened biological sludge flow, enter 'no' in Cell B88 and then enter your thickened biological sludge flow rate in Cell B89 (be sure to use cubic meters per day) and the percent of your thickened sludge total solids that are volatile in Cell B91.

**Suggestion:** If you know your thickened biological sludge flow rate, we highly recommend that you enter 'no' in Cell B88 and enter your own flow rate in Cell B89. The results of the spreadsheet are sensitive to sludge flow rates so these numbers should be as accurate as possible.

6.3.5.3 Rows 94-100

**Do not modify these rows.** These rows contain information critical for the rest of the spreadsheet. You may wish to look at the results and if something does not look correct, recheck your entries in the earlier rows but do not change anything in these rows!
6.3.6 Digester Volume, Biogas Production and Digester Type

The following sections determine the digester volume, biogas production, and digester dimensions. The purposes of these sections are 1) to provide information for the budget-level capital and operating cost estimates for the digesters, 2) to provide information to facilitate estimates of heat loss through the digesters, and 3) to provide the biogas production value needed for numerous other calculations. These sections are not intended to be used to design digesters. Indeed, you may find that the volumes and dimensions calculated in the spreadsheet differ somewhat from digesters at your facility, although they should be similar. The notes below will point out which cells can be compared to your facility and will also provide suggestions to minimize differences between the spreadsheet-calculated results and your actual facility information.

6.3.6.1 Row 103: Digester Working Volume

Do not modify this row, especially Cell B103. Cell B103 calculates the total working volume of the digesters. There is no opportunity for you to input your actual total digester volume because once the digester SRT and sludge flow volumes were specified (as they were earlier in the spreadsheet), digester volume is no longer an independent variable. Therefore, the spreadsheet calculates the volume and reports it here. Do not worry about how well this particular number compares to your actual facility. There are other values later that can be more readily compared.

6.3.6.2 Rows 105-109: Digester Total Volume

Do not modify Rows 107 or 109. In Cell B105, enter 'yes' if you would like to use the default headspace volume for the digesters (10%) and then skip Row 106. If you would like to enter your own headspace volume, enter 'no' in Cell B105 and then enter the digester headspace volume as a percent of the total digester volume in Cell B106.

Row 107 reports the value the spreadsheet will use as the headspace volume based on your selections in Rows 105-106 while Row 109 reports the total digester volume, working volume plus headspace volume.

6.3.6.3 Rows 111-113: Organic Loading Rate and Biogas Production Rate

Do not modify Rows 111 or 113. Row 111 reports the VS loading rate for the total digester volume while Row 113 reports the biogas production rate. You may wish to compare these values to those for your facility. If you have chosen defaults earlier in the spreadsheet and there are significant differences in the VS loading rate and biogas production rate from your facility, you may wish to enter %VSR and sludge flows directly, instead of choosing the defaults. If you have already entered default values and you see significant differences at this point, you should carefully check the values you have entered. In particular, the biogas flow rate is difficult to measure.
6.3.6.4  **Rows 115-118: Number and Type of Digesters**

In Cell B115, enter the total number of digesters. Then examine the result in Cell B116, which reports the volume per digester (in cubic meters). The spreadsheet assumes that all digesters are the same volume. If this is not true at your facility, compare the average volume of your digesters to Cell B116.

If the individual digester volume reported in Cell B116 does not agree with your average volume, you may adjust it by making the following changes *earlier* in the spreadsheet (but not directly in Cell B116): 1) Adjust the number of digesters, 2) Adjust the digester headspace volume, 3) Adjust the sludge flows or 4) Adjust the SRT. From the standpoint of affecting the overall results, many results in the spreadsheet will be sensitive to sludge flow rates and digester SRT. Digester headspace volume will only affect digester capital and operating cost estimates while changing the number digesters will not significantly affect the key results of the spreadsheet.

In Cell B118 enter 'conventional' or 'egg' to describe the geometry of your digesters.

6.3.6.5  **Rows 120-136: Digester Dimensions**

**Do not modify the results in Rows 123, 126, 127, 128, 133, 135 or 136.** Rows 122-128 will show text only if 'conventional' is chosen in Cell B118. Rows 131-137 will show text only if 'egg' is chosen in Cell B118.

If 'conventional' is chosen in Cell B118, enter in Row 124 either 'yes' to select the default digester diameter or 'no' to enter your own digester diameter. You only need to enter this result in the column corresponding to your digester type. Row 122 will show 'yes' column under the digester type that you selected earlier.

If you selected 'no' for Row 124, enter your digester diameter (in meters) in Row 125 in the appropriate column. If you selected 'yes' for Row 124, you may skip Row 125. The digester dimensions are used in the spreadsheet solely to estimate heat loss from the digesters.

If 'egg' is chosen in Cell B118, enter the maximum height of your digester (in meters) in Row 134 in the appropriate column. The digester dimensions are used in the spreadsheet solely to estimate heat loss from the digesters.

6.3.7  **Biogas Composition, Heat Exchange and Temperatures**

6.3.7.1  **Rows 139-142: Biogas Composition**

**Do not modify Cell B140.** Enter the methane percentage in the biogas in Cell B139. This is typically 60-70%. Because the energy content of the biogas is in the methane, this percentage will affect the overall results.
The carbon dioxide percentage in Cell B140 is calculated by difference from the methane percentage, and is not an entered value.

Enter the hydrogen sulfide and siloxane concentrations (parts per million by volume) in Cells B141 and B142, respectively. These numbers are important for selecting gas pretreatment technologies and for estimating sulfur dioxide emissions. They have no impact on the energy content of the biogas.

6.3.7.2 Rows 144-152: Heat Exchange

Do not modify Rows 144, and 147-152. These rows provide the information needed for the heating energy calculations conducted later.

Enter a heat exchanger efficiency (percent) in Cell B146. A reasonable default value is 80%.

6.3.7.3 Rows 154-159: Ambient and Sludge Temperatures

Do not modify Row 159. Enter in Cell B155 the two-letter abbreviation for the State or Province in which your facility is located. Notice in Cell C155 that the average annual ambient temperature (in °C) for your State or Province is shown. This is determined from the Default Temp sheet.

In Cell B156, enter 'yes' if you want to use the default temperature shown in Cell C155 and 'no' if you want to enter your own ambient temperature. If you enter 'no', then enter the ambient temperature (in °C) that you would like to use in Cell B157.

Note: The spreadsheet is set up to handle only one ambient temperature at a time. The default temperatures are average annual values. You may be interested in comparing winter heating requirements versus summer heating requirements because more of the biogas will be needed to heat the digesters in the winter than in the summer. To do this, conduct your analysis twice, once entering the winter ambient temperature for your facility (and then recording the information of interest in the spreadsheet) and then again entering the summer ambient temperature for your facility.

Row 159 shows the estimated temperature of the thickened sludge calculated from the ambient temperature.

6.3.8 Digester Heating Energy

The amount of thermal energy needed to both heat the incoming sludge and to account for losses through the digester walls is estimated in this section. Later in the spreadsheet, you
will be able to compare the thermal energy produced with this estimated requirement to
determine whether you have sufficient heating capacity.

6.3.8.1 Rows 162-173

**Do not modify these rows.** They re-state information from earlier in the spreadsheet for ready calculation and conduct some additional calculations needed to estimate heat losses from the digesters.

6.3.8.2 Rows 174-175: Digester Temperature

In Row 174 under the appropriate column (Row 161 summarizes the digester type for ready reference), enter the temperature (in °C) that is used in the first stage digester. This should be approximately 35-37°C for mesophilic digesters and approximately 55°C for thermophilic digesters. If you are considering a two-stage digester system, enter the temperature of the second stage in Row 175.

6.3.8.3 Rows 176-192

**Do not modify these rows.** These rows conduct the calculations needed to estimate the thermal energy requirements. There are three cells that provide information of particular interest. Cells B181 and B183 show the percentage of the total thermal energy requirement that is due to heat loss through the digester as opposed to the thermal energy requirement required to heat the sludge to digestion temperature. This percentage should be below 30%, and could be as low as 5%. Most of the thermal energy requirement is to heat the sludge.

Cell B192 shows the total thermal energy requirement to heat the digesters (MJ/day). If you would like to quickly see the result of changing the ambient temperature in Cell B157, look in this cell.

**Note:** The spreadsheet does not consider reductions in heat loss from the digesters due to insulation. It also does not consider reductions in thermal energy requirement due to pre-heating the incoming sludge with digested sludge (something that should be considered for thermophilic digesters in particular). To simulate these reductions in the spreadsheet, adjust the ambient temperature in Cell B157. Note that the spreadsheet will not automatically account for additional capital and operating costs due to these reductions but that can be adjusted directly in the Economics worksheet (discussed later in this manual).
6.3.9 Digester Mixing Energy

6.3.9.1 Rows 195-196: Selecting Mixing Status
Enter 'yes' or 'no' in the appropriate cells of these rows to indicate whether the digester is mixed or not. The mixing energy is used in the Economics worksheet to estimate the electricity energy costs to operate the anaerobic digesters.

6.3.9.2 Rows 197-207
Do not modify these rows. The results shown in Cells B205 and B207 show the electrical energy estimates for mixing of the digesters in Watts and MJ/day, respectively.

6.3.10 Energy Recovery Systems

6.3.10.1 Row 210: Selecting Energy Recovery System
The spreadsheet considers 10 different energy recovery systems described in Table 2-2. Enter 'yes' in the appropriate cells for every energy recovery system to be operated simultaneously at your facility. From a practical standpoint, we recommend selecting either a boiler, or one electricity producing system, or a boiler plus one electricity producing system. Nevertheless, many electricity producing systems may be simultaneously selected.

6.3.10.2 Row 211: Select Individual Boiler Capacity
If you selected 'yes' in Cell B210 for boiler, enter the individual boiler capacity (horsepower) in Cell B211. If you are not selecting a boiler, you may leave Cell B211 blank.

6.3.10.3 Row 212
Do not modify this row. It reports the capacity of one boiler in MJ/day.

6.3.10.4 Row 213: Maximum Possible Capacity
Do not modify this row. This row is provided for information to help you select possible energy recovery systems. The spreadsheet has calculated the total possible usable energy that could be produced if all the biogas produced was used for a given energy recovery technology. For example, if all the biogas was used to operate an engine cogeneration system, Cell C213 shows the maximum power output, as electricity, that could be achieved for the biogas production shown in Cell B113, assuming the specified methane concentration and the production efficiency for the generation system. The results in this row will not be accurate until you have entered the production efficiencies in Rows 218-226 (see below).

6.3.10.5 Rows 214-216: Select Individual Capacities of Systems and Numbers
Do not modify Row 216. In Row 214, select the capacity (kW) of the individual energy recovery unit. Do not modify Cell B214, which simply shows the capacity of the boiler selected earlier in kW units. In Row 215, select the number of individual energy recovery units for each.
Table 6-2. Energy Recovery Systems.

<table>
<thead>
<tr>
<th>Abbreviation in the Spreadsheet</th>
<th>Technology</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler</td>
<td>Standard boiler. Can use natural gas or biogas.</td>
</tr>
<tr>
<td>Engine-Co</td>
<td>Engine Cogeneration: Internal combustion engine to produce electricity. Waste heat can be used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>Steam Turb</td>
<td>Steam Turbine: Boiler produces steam used to turn turbines for electricity generation. Steam also used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>Gas Turb</td>
<td>Gas Turbine: Gas combustion produces exhaust gases to turn turbines for electricity generation. Waste heat can be used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>Microturb</td>
<td>Microturbine: Conceptually the same as gas turbines, only much smaller and less efficient. Waste heat can be used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>MC Fuel Cell</td>
<td>Molten Carbonate Fuel Cell. Second generation fuel cell used for electricity generation. Still being tested with biogas. Waste heat can be used for thermal energy.</td>
</tr>
<tr>
<td>PA Fuel Cell</td>
<td>Phosphoric Acid Fuel Cell. First generation fuel cell used for electricity generation. Being used with biogas. Waste heat can be used for thermal energy.</td>
</tr>
<tr>
<td>Stirling Eng</td>
<td>Stirling Engine. External combustion engine to produce electricity. Waste heat can be used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>Direct Drive</td>
<td>Engine: Internal combustion engine to produce mechanical force. Waste heat can be used for thermal energy (cogeneration).</td>
</tr>
<tr>
<td>Combined Cycle</td>
<td>Combined Cycle Turbine: Two stage turbine process that maximizes electricity production from steam. Waste heat can also be used for thermal energy (cogeneration).</td>
</tr>
</tbody>
</table>

Instead, compare the total capacity of the energy recovery systems that you have selected (which is calculated in Row 216) to the maximum possible capacity shown in Row 213. If the value in Row 216 exceeds that in Row 213, you will have to supplement your energy recovery systems with natural gas.

6.3.10.6 Rows 218-226: Production Efficiencies

*Do not modify Rows 220, 223 or 226.* The results in these rows are based on your input and will be used in subsequent calculations.
In Rows 218, 221 and 224, select 'yes' if you would like to use the default production efficiencies for thermal energy production, electricity production and mechanical energy production, respectively. If you want to enter your own production efficiencies, select 'no' in Rows 218, 221 and 224 and then enter the appropriate efficiency values in Row 219 (for thermal energy production), Row 222 (for electricity production) and Row 225 (for mechanical energy production).

Note that you should not enter electricity production efficiencies for boilers (Column B) or Direct Drives (Column J). Also, enter mechanical energy production efficiencies only in Column J (for Direct Drives) and not in any of the other columns. Finally, you need only complete the cells in Rows 218, 219, 221, 222, 224 and 225 for the energy recovery systems that you selected in Row 210.

6.3.11 Biogas Needs and Natural Gas Supplementation

6.3.11.1 Rows 228-245

Do not modify Rows 228-235, 237, 239-242 or 244-245. Rows 228-245 provide the calculations for comparing the biogas produced to the amount required for the energy recovery systems that you have selected.

Rows 228 and 229 calculate the biogas required for the individual energy recovery systems selected. The total biogas required is summed in Cell B230. Cell B231 then estimates the amount of biogas that could be used without purification to remove hydrogen sulfide or siloxanes. We have assumed that only boilers, steam turbines and Stirling engines can be effectively operated without purification. The rest of the energy recovery systems should have gas pretreatment to ensure effective operation. The amount of biogas requiring purification is estimated in Cell B232 and the percent of the total biogas that should be purified is indicated in Cell B233.

Row 234 indicates whether you have used all your biogas or not. If you have not used all the biogas, Cell B234 will show a percentage and Cell C234 will state ‘You still have some biogas not used’. If you have used more biogas than is produced, Cell B234 will state ‘Biogas needed exceeds Biogas production, % is’ with Cell C234 then showing the percentage of biogas used. This number will be greater than 100% in this case. Using the information in Row 234, you may wish to adjust your selection of energy recovery systems to better utilize biogas.

If Cell B235 is positive, you have excess biogas that can be sent offsite. The amount that you send offsite can be specified in Cell B236. If Cell B235 is negative, you cannot send biogas offsite. Cell C236 states whether the amount you select in Cell B236 is appropriate. Any excess biogas that must be flared is shown in Cell B237.
Enter 'no' in Cell B238 (our recommended default value) if you either have no biogas to be sent offsite or you do not wish to purify it. Entering 'yes' automatically adjusts the responses in Rows 239 and 240 and will affect the gas pretreatment costs and the emissions, but not the energy recovery performance.

Cell B243 allows you to enter the amount of natural gas to be purchased (cubic meters per day) to fully utilize the energy recovery systems selected. Cell B241 tells you whether natural gas is needed or not, based on the expected biogas utilization for the energy recovery systems selected while Cell B242 estimates the amount of natural gas needed. This value could be entered into Cell B243. Cells B244 and B245 provide additional information about your natural gas selection and the subsequent energy recovery calculations.

**Note:** The spreadsheet calculates the energy produced from your selections in Rows 210-216, not from the amount of biogas produced. Therefore, if you enter a value in Cell B243 that is less than that suggested in Cell B242, the energy recovery calculations conducted later will be inaccurate. You should examine Rows 247-250 first, however, before selecting the amount of natural gas needed because any changes that you make from considering the thermal energy balance will affect the amount of natural gas that may be needed.

### 6.3.11.2 Rows 247-250: Thermal Energy Balance

**Do not modify these rows.** These rows provide information about how much thermal energy is produced compared to how much is needed to heat the sludge and digesters.

**Insufficient thermal energy produced.** If the amount of thermal energy produced by the energy recovery systems that you have selected is less than that needed to heat the sludge and digesters, you will see the statement 'Balance Not Ok' in Cell B247. You will also see the statement 'Demand Exceeds Production' in Cell B249 with an estimate of how short you are (in percentage) in Cell C249. Cell B250 will suggest that 'higher ERS capacity or unit number may be required.' So return to Rows 210-216 and make appropriate adjustments (e.g. increase the number of boilers).

**Excess thermal energy produced.** If the amount of thermal energy produced by the energy recovery systems is more than that needed to heat the sludge and digesters, you will see the statement 'Balance Not Ok' in Cell B247. You will also see the statement 'Production exceeds Demand' in Cell B248 with an estimate of the excess thermal energy produced (in percentage) in Cell C248. You do not have to take any action, but you may wish to change your selection of energy recovery systems. For example, if you have excess thermal energy production, you could decrease the number of boilers selected and increase the number of electricity generating systems used to decrease the ‘waste’ thermal energy produced.
**Thermal energy balanced.** If the amount of thermal energy produced by the selected energy recovery systems exactly equals the amount needed, Cell B247 will report 'Balance OK.' Because this is practically impossible, you will likely never see this statement. We recommend adjusting your energy recovery selections to provide a slight excess of thermal energy above what is needed to heat the sludge and digester.

**Note:** After making any changes in Rows 210-216 to better balance your thermal energy production with demand, return to Cell B243 and compare it to Cell B242. If, for example, you added additional boilers to meet your thermal energy needs, this may require supplemental natural gas. Cell B241 will indicate the need and Cell B242 will recommend the amount required.

6.3.12 Biogas Pretreatment

In this section of the spreadsheet you will decide whether to implement biogas pretreatment technology. The selection of specific technology occurs in the Economics worksheet.

6.3.12.1 Rows 254-257: Carbon Dioxide Removal

**Do not modify Rows 255 or 257.** None of the energy recovery technologies examined in the spreadsheet require carbon dioxide removal. Biogas sent offsite, however, may require carbon dioxide removal. If you are not planning to send or sell biogas offsite, we recommend entering 'no' into each of Cells B254-M254. If you are planning to send or sell biogas offsite and its energy content must be improved by removing carbon dioxide, then enter 'yes' into Cells B254 and M254 (off-site) and enter a removal efficiency (percentage) into Cell B256. If you decide to remove carbon dioxide for use in any other energy recovery system, enter 'yes' in Cell B254 and also in the appropriate cell for that energy recovery system. Note, if you have entered 'yes' in any of the Cells C254-M254, you should enter 'yes' in Cell B254 as well.

Cell B257 indicates what the percent carbon dioxide will be in the purified biogas based on your entry in Cell B256.

6.3.12.2 Rows 259-262: Hydrogen Sulfide Removal

**Do not modify Rows 260 or 262.** We highly recommend hydrogen sulfide removal for all technologies although the amount of hydrogen sulfide removal required depends on the energy recovery technology selected (see the full report for detail). Therefore, we recommend entering 'yes' in Cell B259 and 'yes' for each energy recovery technology being used. You will notice in Row 253 directly under the energy recovery technology heading a statement 'Decision required' for each energy recovery technology that you have selected in Rows 210-216. This is a reminder of which cells in Row 259 should be completed. Note that entering 'yes' or 'no' for an energy recovery technology that is not selected will have no impact on the spreadsheet.
calculations so for simplicity, you may wish to enter the same response into each cell of Row 259.

If you enter 'yes' into any cells of Row 259, enter the removal efficiency desired (in percent) in Cell B261. The remaining hydrogen sulfide concentration (in ppm) is then shown in Cell B262 based on the input concentration that you entered earlier and the removal efficiency. We recommend checking the full report to identify the remaining hydrogen sulfide concentration allowed for your selected energy recovery technology and adjusting the removal efficiency accordingly.

6.3.12.3 Rows 264-267: Siloxane Removal

Do not modify Rows 265 or 267. We highly recommend siloxane removal for all technologies although the amount of siloxane removal required depends on the energy recovery technology selected (see the full report for detail). Therefore, we recommend entering 'yes' in Cell B264 and 'yes' for each energy recovery technology being used. You will notice in Row 253 directly under the energy recovery technology heading a statement 'Decision required' for each energy recovery technology that you have selected in Rows 210-216. This is a reminder of which cells in Row 264 should be completed. Note that entering 'yes' or 'no' for an energy recovery technology that is not selected will have no impact on the spreadsheet calculations so for simplicity, you may wish to enter the same response into each cell of Row 264.

If you enter 'yes' into any cells of Row 264, enter the removal efficiency desired (in percent) in Cell B266. The remaining siloxane concentration (in ppm) is then shown in Cell B262 based on the input concentration that you entered earlier and the removal efficiency. We recommend checking the full report to identify the remaining siloxane concentration allowed for your selected energy recovery technology and adjusting the removal efficiency accordingly.

6.3.13 Biogas Volume and Composition after Purification

6.3.13.1 Rows 271-280

Do not modify these rows. These rows conduct intermediate calculations that will be used subsequently. There is no need for additional input, nor will you likely need the information contained in these rows.

6.3.14 Energy Produced from Energy Recovery Systems

Do not modify Rows 283-309. These rows conduct and summarize the energy calculations for the selected energy recovery technologies. Although they should not be modified and no additional input is required, you will find information of interest in these rows as described below.
6.3.14.1 Rows 293-298: Energy Recovered Per System

Rows 293, 295 and 297 show the potential energy (MJ/day) that can be recovered as thermal energy, electricity and mechanical energy, respectively, given your selections earlier in the spreadsheet. Rows 294, 296 and 298 show the actual energy recovered (MJ/day) as thermal energy, electricity and mechanical energy, respectively. The difference between the two rows is a logic step only; any efficiency differences were accounted for earlier in the spreadsheet. By looking at Rows 294, 296 and 298 you can see how much and what form of energy is produced by each energy recovery system selected.

6.3.14.2 Rows 301-309: Total and Net Energy Recovered

The total amounts of thermal energy, electricity and mechanical energy recovered are shown in Rows 302-304 while the net amounts of energy recovered are shown in Rows 307-309. Only thermal energy recovered changes. In Row 307, the net thermal energy recovered is the difference between the thermal energy produced, shown in Cell B302 and the thermal energy required to heat the sludge and digester (the heating requirement is shown in Cell B192). If the net thermal energy recovered in Cell B307 is negative, return to Rows 247-250 for information about how to change this.

6.3.15 Greenhouse Gas and Air Emissions

The remaining rows in the Technical worksheet are used to estimate emissions due to the utilization of biogas onsite as well as the reduction in emissions that occur because renewable fuel is being used.

6.3.15.1 Rows 312-313: Combustion Efficiencies

In Row 312, enter 'yes' to choose the default combustion efficiency (95%) for every energy recovery technology that you have selected. If you enter 'yes' in Row 312, you may ignore Row 313. If you choose not to use our default combustion efficiencies, enter 'no' in the appropriate cells of Row 312 and then enter the combustion efficiency (percent) in the appropriate cells.

6.3.15.2 Row 314

Do not modify this row.

6.3.15.3 Rows 315-321: Onsite Emissions per Energy Recovery Technology

Do not modify these rows. These rows show the resulting emissions for 7 air pollutants: carbon dioxide (Row 315), methane (Row 316), sulfur dioxide (Row 317), nitrogen oxides (Row 318), non-methane volatile organic compounds (Row 319), total particulate matter (Row 320) and carbon monoxide (Row 321). The units used are Mg per year, which are equivalent to metric tonnes per year. These estimates are based on mass balances (for carbon dioxide, methane and sulfur dioxide) and emissions factors contained in the ERS Emission Factors worksheet (for NOx, NMVOC, TPM and CO). Although natural gas emissions may be slightly different as
indicated in the ERS Emission Factors worksheet, we have used the biogas emissions factors. This may lead to minor differences in emissions if natural gas is used to supplement the biogas for energy production as discussed earlier.

6.3.15.4 Rows 325-333: Total Onsite Emissions

Do not modify these rows. Row 325 indicates how much of the total gas used to produce energy was biogas. If natural gas is used to supplement the biogas, the amount in Cell B325 will be less than 1.

Row 326 indicates the total carbon dioxide released due solely to biogas conversion and includes the carbon dioxide that was in the biogas gas before conversion. As indicated in Row 327, however, the only carbon dioxide that counts as an emission for greenhouse gas accounting purposes (in accordance with the protocol used by the Intergovernmental Panel on Climate Change) is the carbon dioxide produced from the combustion of supplemental natural gas. Biogas is considered a renewable, biomass-based fuel. Therefore, if only biogas is used at your facility, the entry in Cell B327 will be zero.

6.3.15.5 Rows 337-339: Offset Emissions

Do not modify these rows. If you have produced net electricity as indicated in Cell B308, you will require less electricity from the grid and therefore decrease the emissions arising from the production of grid electricity. These rows estimate the emissions reductions occurring (the emission that are offset because that electricity production is no longer needed) due to your onsite generation of electricity.

There are two points to consider with respect to these offset emissions. First, the emissions generated due to the production of grid electricity depend on the specific electricity production technologies used. If you look at the Default Temp worksheet, you will see the emissions values by State and Province. For example, if you are in a jurisdiction that depends primarily on coal-fired electricity generation, the emissions per kWh of produced electricity will be greater than those in a jurisdiction with more hydroelectric generation.

Second, the total emissions reduction due to your use of biogas to produce electricity is equal to your on-site emissions minus the offset emissions. If that difference is negative, then your facility is reducing emissions overall. This will likely be true for carbon dioxide in virtually all cases. On the other hand, this may not be true for sulfur dioxide, for example, if your biogas has a high hydrogen sulfide concentration and is not pretreated. Similarly, your total NOx emissions may also increase, depending on the jurisdiction in which you are located. Note that the total emissions reduction may have little impact on your ability to obtain air permits, however, which will likely be determined solely on your onsite emissions.
The offset emissions are based solely on onsite electricity generation. This provides an underestimate of the true offset emissions if you have chosen mechanical energy production using direct drives.

6.4 Economics Worksheet

6.4.1 Using the Economics Worksheet

Once you have completed the Technical worksheet, you are ready to complete the Economics worksheet. We recommend working through the entire Economics worksheet in the order that follows, even if you are not interested in economic estimates for a particular technology. For example, you may not be interested in costs for anaerobic digester construction and operation. That is fine, just choose default values for those technologies you are not interested in (you may also choose default values for those you are interested in). By doing this, you will ensure that the spreadsheet has all the appropriate information. You will also see what is being calculated throughout the spreadsheet in case you decide later that you are interested in a specific set of economic calculations. Then, at the bottom of the worksheet, we identify options to include or exclude different technologies in your final economic calculations. So while it may be tempting to skip right to the bottom, doing so may jeopardize the accuracy of the results.

6.4.1.1 Rows 1-9

The first 9 rows provide the color-coded legend. Of the 8 colors shown, the one that matters most to you is light-green (this may also look light-blue). As was the case with the Technical worksheet, cells that are labeled this color require input from you. The rest of the colors were more useful to us when we prepared the spreadsheet, but they also serve to help break up the monotony of row upon row of numbers.

6.4.2 Anaerobic Digester Capital Costs

6.4.2.1 Row 12: Choosing Digester Material and Shape

The capital cost estimates for the anaerobic digesters depend on the material and shape chosen. Look at (but do not modify!) the result in Cell C12. If it says 'yes' then you chose a conventional digester shape in Cell B118 in the Technical worksheet. In that case, you must enter 'no' in each of Cells D12 and E12. If that is not what you want, go back to the Technical worksheet starting at Cell B118 and make the appropriate changes.

If Cell C12 says 'no', then you chose an egg-shaped digester in the Technical worksheet. In that case, enter 'yes' into either Cell D12 or Cell E12, but not both. Enter 'no' into the other one of those cells.

6.4.2.2 Rows 13-16: Digester Unit Capital Cost

In Cell B13, enter 'yes' if you want to use our default unit capital cost for anaerobic digester construction. If you want to enter your own default unit capital cost, enter 'no' in Cell
B13 and then enter the amount ($ per cubic meter volume) into either Cell C15, D15 or E15, depending on which digester type and material is used.

Do not modify any of the other cells in these rows. Cell B16 then shows the unit capital cost that will be used for the calculation, based on your selections.

6.4.2.3 Row 17: Capital Cost of Anaerobic Digesters

Do not modify this row. Cell B17 shows the capital cost for constructing the anaerobic digester volume determined from your entries in the Technical worksheet. This number can be compared to other capital cost estimates to determine the relative costs of the different technologies being considered.

6.4.3 Energy Recovery System Capital Costs

6.4.3.1 Rows 21-22

Do not modify these rows. They are bringing over information from the Technical worksheet.

6.4.3.2 Rows 23-25: Energy Recovery System Unit Capital Cost

In Cell B23, enter 'yes' if you would like to use our default unit capital costs. If you would like to enter your own unit capital costs, enter 'no' in Cell B23 and then enter the appropriate unit capital cost (in $/kW) in Cells C25-L25. You need only enter values for the energy recovery systems that are being considered. If you have forgotten which those are, look at the entries in Row 22. Those energy recovery systems that you selected in the Technical worksheet are noted 'yes' in Row 22. Do not modify any other cells in these rows.

6.4.3.3 Rows 26-27

Do not modify these rows. These rows conduct logic calculations.

6.4.3.4 Rows 28-29: Energy Recovery System Capital Costs

Do not modify these rows. Row 28 contains the capital cost estimates for each energy recovery system that you have selected. Cell B29 then contains the total capital cost of all the energy recovery systems you have chosen.

6.4.4 Gas Pretreatment Systems Capital Costs

6.4.4.1 Row 33: Choosing Gas Pretreatment System

In Row 33, indicate by entering 'yes' into the appropriate cell which gas pretreatment system you would like. We have provided 6 alternatives: iron salts addition to the sludge, iron sponge, Sulfa Treat, Binax, Apollo and activated carbon. For information about the performance of these technologies, please see the report. If you want to choose a different gas pretreatment...
technology, then enter 'yes' in Cell D33 and 'no' in the other cells. Although called Iron Sponge, Cell D33 also serves as the 'Other Technology' cell.

**Note:** Do not modify Cells F33 or H33. If you chose to remove carbon dioxide in the Technical worksheet (Cell B254), then Binax will be selected and you will see 'yes' in Cell F33. If you chose to remove siloxane in the Technical worksheet (Cell B264), then activated carbon will be selected and you will see 'yes' in Cell H33.

**6.4.4.2 Rows 34-36: Gas Pretreatment System Unit Capital Cost**

To use our default unit capital costs, enter 'yes' in Cell B34. If you have chosen the 'Other Technology' option noted above or otherwise do not want to use the default values, enter 'no' in Cell B34.

If you have entered 'no' in Cell B34, then you must enter unit capital costs into the appropriate cells of Row 36. Note that the units are $ per cubic meter of biogas treated per day except for iron salts addition which has units of $ per Mg (metric tonne) of volatile solids per day. If you have chosen the 'Other Technology' option, enter the unit capital cost ($ per cubic meter of biogas treated per day) for your technology in Cell D36.

**6.4.4.3 Rows 37-38**

Do not modify these rows. These rows conduct logic calculations.

**6.4.4.4 Rows 39-40: Gas Pretreatment System Capital Costs**

Do not modify these rows. Row 39 contains the capital cost estimates for each gas pretreatment system that you have selected. Cell B40 then contains the total capital cost of all the gas pretreatment systems you have chosen.

**6.4.5 Replacement Costs for Gas Pretreatment and Energy Recovery Systems**

The full life cycle cost of a decision should include the expected replacement costs for components that may need replacing during the total evaluation time. The life cycle cost calculations conducted at the bottom of the worksheet will be based on a planning period that you will specify in Cell B80.

**Suggestion:** We recommend that you jump down to Cell B80 now and enter your desired planning cycle time (we recommend 20 years) and then use that to think about your entries for the replacement cost calculations.

**6.4.5.1 Row 43: Discount Rate**

Enter in Cell B43 the value of the discount rate (percent) that you wish to use for all subsequent calculations.
Note: Although located here, this entry is key for many other calculations later in the worksheet. We recommend entering a value of 5% as a default. For more discussion on the role of the discount rate in the life cycle calculations please see the full report.

6.4.5.2 Row 44: Service Life of Gas Pretreatment System
Enter in Cell B44 the service life (years) expected for the gas pretreatment systems used.

6.4.5.3 Row 45: Number of Replacements Required for Gas Pretreatment System
Enter in Cell B45 the number of full replacements that you expect to make of the gas pretreatment system over the course of the planning period specified in Cell B80. This is simply determined as the planning period divided by the service life entered in Cell B44. If the service life is greater than or equal to the planning period, enter 0 in Cell B45. Do not enter a number greater than 4 in Cell B45 (the economic calculations have not been designed for greater than 4 replacements during a planning period).

6.4.5.4 Rows 46-47: Replacement Costs for Gas Pretreatment Systems
Do not modify these rows. Row 46 shows the replacement cost for each gas pretreatment system selected in today's dollars. This incorporates the time value of money for expenditures made in the future. The specific time value of money calculations are conducted in the “Factor-ERS” worksheet.

6.4.5.5 Rows 48-49: Capital and Replacement Costs for Gas Pretreatment Systems
Do not modify these rows. Row 48 shows the sum for each gas pretreatment system of the initial capital and replacement costs in today's dollars. Cell B49 shows the total sum of the initial capital and replacement costs for all of the gas pretreatment systems in today's dollars. If there are no replacement costs, the value in Cell B49 should be equal to the value in Cell B40 (just initial capital costs).

6.4.5.6 Row 52
Do not modify this row. It contains the discount rate that you entered in Cell B43.

6.4.5.7 Rows 53-54: Service Lives for Energy Recovery Systems
In Cell B53, enter 'yes' if you would like to use our default service lives for the energy recovery systems. You can see our default values in the other cells of Row 53. If you enter 'no' in Cell B53, then enter the service lives you prefer in the appropriate cells (C54-L54) of Row 54.

6.4.5.8 Row 55
Do not modify this row. It contains logic calculations.
6.4.5.9 Row 56: Number of Replacements required for Energy Recovery Systems

Enter in the appropriate cells the number of times each energy recovery system must be replaced. Keep in mind the planning period that you specified in Cell B80. For example, if you are using boilers and the service life of a boiler is taken to be 15 years, then you will have to replace the boiler once during a 20-year planning period and should enter 1 into Cell C56. If the planning period is less than or equal to the service lives of the energy recovery equipment, then enter 0 in the appropriate cells of Row 56.

6.4.5.10 Row 57: Replacement Costs for Energy Recovery Systems

Do not modify this row. Row 57 shows the replacement cost for each energy recovery system selected in today's dollars. This incorporates the time value of money for expenditures made in the future. The specific time value of money calculations are conducted in the FactorERS worksheet.

6.4.5.11 Rows 58-61: Capital and Replacement Costs for Energy Recovery Systems

Do not modify these rows. Rows 58 and 60 shows the sum for each energy recovery system of the initial capital and replacement costs in today's dollars (the difference is the logic step used in Row 59). Cell B61 shows the total sum of the initial capital and replacement costs for all of the energy recovery systems in today's dollars. If there are no replacement costs, the value in Cell B61 should be equal to the value in Cell B29 (just initial capital costs).

6.4.6 Plant Electricity Use and Bioelectricity Production

This part of the spreadsheet estimates the total electricity needed by your treatment facility (for all operations, not just the anaerobic digestion and biogas processes) and compares that to the amount of bioelectricity produced from biogas and energy recovery. The results of these calculations are then used later for estimate savings from reduced electricity use and possible revenues from the sale of excess electricity.

6.4.6.1 Rows 64-66: Offsite Electricity Demand

Do not modify Cell B66. In Cell B64, enter 'yes' is you would like to use our default electricity demand. We are estimating that conservatively as 0.4 kWh of electricity required per cubic meter of wastewater treated. This is for operation of the entire facility. We have used the term 'offsite electricity demand' to indicate that this is electricity you would normally have to purchase from the grid if you did not produce bioelectricity onsite.

If you enter 'no' in Cell B64, enter your annual electricity requirement for the entire facility in Cell B65 (kWh/year). If you would like to compare your value to our estimate, simply type 'yes' in Cell B64 and our estimate will appear in Cell B66. Then to switch back to your number, type 'no' in Cell B64 and Cell B66 will show the number you entered in Cell B65.
6.4.6.2 Row 67

Do not modify this row. Cell B67 shows the onsite electricity production predicted from the Technical worksheet.

6.4.6.3 Rows 68-72: Electricity Required for Anaerobic Digestion

Do not modify Rows 70-72. Enter 'yes' in Cell B68 to use our default estimate of the electricity required to mix the anaerobic digesters. Our estimate is shown in Cell C68 (brought over from the Technical worksheet). If you enter 'no' in Cell B68, then enter your estimate of the electricity required for anaerobic digester operation in Cell B69. Because our default estimate is for mixing only and does not include savings due to egg-shape digesters or energy requirements for pumping (e.g. for heating), we recommend that if possible, you should use your estimates here.

Row 70 conducts a logic calculation and Row 71 calculates for your information the percentage of the total electricity produced onsite that would be required just for operation of the anaerobic digesters. Row 72 shows the bioelectricity remaining to be used for other operations at the facility.

6.4.6.4 Rows 73-75: Bioelectricity Availability

Do not modify Rows 74-75. In Cell B73, enter 'yes' if you would like to use the bioelectricity produced onsite to help meet the facility's electricity demand. We recommend entering 'yes' because the cost savings of offsetting electricity will likely be greater than the revenue generated if this electricity was sold to the grid.

Cell B74 shows how much surplus bioelectricity is available to be sold offsite. If you enter 'no' in Cell B73, then all of the produced bioelectricity can be sold. If you enter 'yes' in Cell B73, then only the electricity that is not consumed at your facility will be available to be sold offsite. Unless you have chosen to augment your biogas with natural gas (see the Technical worksheet), you will likely not have excess electricity beyond your plant needs and should see a negative number in Cell B74. There may be exceptions, however, if for example you are using an advanced process to greatly increase your VSR (e.g. sonification) or your plant electricity requirements are relatively low (e.g. in the vicinity of 0.2 kWh per cubic meter treated). Cell B75 shows the amount of bioelectricity produced onsite that will be used to meet onsite demand.

6.4.7 Salvage Values (and Planning Period)

6.4.7.1 Row 80: Planning Period

If you have not already done so, enter the planning period of interest (years) in Cell B80. Although primarily used for calculating the salvage values, the planning period entered here is also used at the bottom of the spreadsheet to conduct the life cycle cost calculations. Therefore, a value must be entered in Cell B80. We recommend 20 years.
6.4.7.2 Rows 78-79, 81-96: Salvage Value Calculations

Do not modify these rows. These rows determine the salvage values at the end of the planning period for the energy recovery systems and the anaerobic digester. Row 85 indicates the salvage values for each energy recovery system selected (in today's dollars) and Cell B86 shows the total salvage value for all of the energy recovery system equipment (in today's dollars). Cell B91 shows the salvage value for the anaerobic digesters (in today's dollars). Rows 95 and 96 show the salvage values for the energy recovery system and anaerobic digesters, respectively, on an annualized basis.

6.4.8 Revenues

6.4.8.1 Rows 99-100: Excess Heat

Row 100 asks you to determine whether you will sell the excess thermal energy shown in Row 99. Enter 'yes' in Cell B100 if you plan to sell heat and enter 'no' otherwise. Note that the spreadsheet does not include the capital and operating costs for installing the capability to sell thermal energy offsite. Therefore, unless you already have such capability at your facility, we recommend that you enter 'no' in Cell B100.

6.4.8.2 Rows 102-113: Setting Prices and Calculating Revenues

The spreadsheet assumes that you have three possible products to sell: electricity, heat and biogas. If you do not plan to sell products offsite, then enter 'no' in Cell B103 and enter 0 in Cell C104. Also enter 'no' in each of Cells B106 and B109. Then check Cell B113, which shows your net annual revenue, to ensure that it is $0. If Cell B113 is $0, continue. If Cell B113 is not $0 then check the following: Cell B100 (decision on selling excess heat) should be set to 'no'; and Cell B236 in the Technical worksheet should be set to 0.

If you plan to sell products offsite, do the following. To sell electricity, heat and biogas using the default price, enter 'yes' in Cells B103, B106 and/or B109 respectively. We do not recommend using our default prices, however. You should use an appropriate price for your facility. To enter your own prices for electricity, heat and biogas, enter 'no' in Cells B103, B106 and/or B109, respectively. Then enter the electricity price ($/kWh) in Cell C104, the heat price ($/MJ) in Cell C107 and/or the biogas price ($/cubic meter) in Cell C110.

6.4.9 Savings

The spreadsheet assumes that you could achieve savings in electricity and natural gas consumption by using biogas to generate electricity and thermal energy. This section allows you to set the prices from which the savings will be calculated.
6.4.9.1 Rows 116-118: Normal Electricity Price

Do not modify Row 118. The spreadsheet allows consideration of two electricity prices: a peak electricity price that might be paid during peak usage times and a normal electricity price that would be paid at all other times. If this is overly simplistic for your particular facility, you will need to define a peak price to use and an average electricity price to use for non-peak times.

If you wish to use our default normal electricity price, enter 'yes' in Cell B116 and skip Row 117. We recommend, instead, that you enter 'no' in Cell B116 and enter the normal, non-peak electricity price that you pay ($/kWh) in Cell C117.

6.4.9.2 Rows 119-121: Peak Electricity Price

Do not modify Row 121. If you wish to use our default peak electricity price, enter 'yes' in Cell B119 and skip Row 120. We recommend, instead, that you enter 'no' in Cell B119 and enter the peak electricity price that you pay ($/kWh) in Cell C120.

6.4.9.3 Rows 122-123: Percentage of Normal versus Peak Price

Do not modify Row 123. Electricity generated and used onsite will offset purchased electricity. In Cell B122, enter the percentage of onsite produced electricity that offsets electricity purchased at the normal price. For example, if you enter 100%, then all of the electricity produced onsite is offsetting only the less expensive normal priced electricity. Alternatively, if you enter 0%, then all of the electricity produced onsite is offsetting the more expensive peak priced electricity.

Cell B123 shows the percentage of onsite produced electricity that is offsetting electricity purchased at the peak price. This is calculated from Cell B122.

6.4.9.4 Row 124: Conversion Efficiency of Electricity to Mechanical Energy

If direct drives are used as an energy recovery system, the electricity saved is equal to the energy produced by the drive and the losses that would be incurred due to inefficiencies of converting electricity to mechanical energy if the drive was otherwise operated with electricity. Entering an efficiency less than 100% in Cell B124 accounts for the extra benefit of using direct drives instead of electricity generators. As a default, we recommend entering 95%.

6.4.9.5 Row 125: Total Electricity Saved

Do not modify this row. Cell B125 shows the total electricity saved (including the mechanical energy if direct drives are used) at your facility.

6.4.9.6 Rows 126-128: Natural Gas Prices

Do not modify Row 128. If you wish to use our default natural gas price, enter 'yes' in Cell B126 and skip Row 127. We recommend, instead, that you enter 'no' in Cell B126 and enter the natural gas price that you pay ($/cubic meter) in Cell C127.
6.4.9.7 Rows 129-130

Do not modify these rows. Row 129 is a conversion factor. Row 130 shows the amount of thermal energy used to heat the sludge and digesters. Note that this is all the thermal energy required. If biogas does not provide sufficient thermal energy to heat the sludge and digesters, this full savings will not be realized, but this is considered separately as an operating cost.

6.4.9.8 Rows 131-133: Annual Savings

Do not modify these rows. Row 131 shows the annual savings due to onsite electricity generation and USge while Row 132 shows the annual savings due to generating thermal energy for heating onsite. Row 133 provides the total annual savings from both of these activities.

6.4.10 Salvage Value as Revenue

The salvage values for the energy recovery systems and anaerobic digester were determined in Rows 79-96. Row 136 asks if you would like the salvage value to be included in the overall calculations to be conducted later. We highly recommend that you answer 'no' in Cell B136 unless you are confident both in the predicted salvage values and in your ability to sell the materials at the end of the planning period. Answering 'yes' in Cell B136 will make the project look economically more attractive than perhaps it should.

6.4.11 Anaerobic Digester Operating and Maintenance Costs

6.4.11.1 Rows 139-140, 142, 144-147

Do not modify these rows. The annual O&M costs are shown in Row 147.

6.4.11.2 Rows 141 and 143: Selecting Unit O&M Costs

If you would like to use our default O&M costs or are not interested in including the anaerobic digester costs, enter 'yes' in Cell B141 and skip Row 143 (the option to not include anaerobic digester costs is provided later in the spreadsheet). If you are interested in including the anaerobic digester O&M costs, you may enter 'yes' in Cell B141 to use our default values. When the anaerobic digester is considered in the complete economic evaluation, however, we have found the results to be very sensitive to the O&M costs. Therefore, if you are including anaerobic digester costs in your evaluation, we recommend that, if possible, you enter 'no' in Cell B141 and then enter the appropriate O&M unit costs ($ per Mg VS) in Row 143 (for the anaerobic digestion technology chosen).

6.4.12 Gas Pretreatment System Operating and Maintenance Costs

6.4.12.1 Rows 150, 152, 154-158

Do not modify these rows. The daily O&M costs for each system used are shown in Row 156 and the overall daily O&M costs are show in Row 157. The annual O&M costs are shown in Row 158.
6.4.12.2 Rows 151 and 153: Selecting Unit O&M Costs

If you would like to use our default O&M costs, enter 'yes' in Cell B151 and skip Row 153. We recommend, however, that if possible you enter your own O&M costs. In that case, enter 'no' in Cell B151 and then enter the appropriate O&M unit costs ($ per cubic meter of biogas treated per day for all but iron salts addition. Use $ per Mg VS for iron salts addition) in Row 153 (for the gas pretreatment technologies chosen).

6.4.13 Energy Recovery System Operating and Maintenance Costs

6.4.13.1 Rows 161, 163, 165-170

Do not modify these rows. The daily O&M costs for each system used are shown in Row 168 and the overall daily O&M costs are show in Row 169. The annual O&M costs are shown in Row 170.

6.4.13.2 Rows 162 and 164: Selecting Unit O&M Costs

If you would like to use our default O&M costs, enter 'yes' in Cell B162 and skip Row 164. We recommend, however, that if possible you enter your own O&M costs. In that case, enter 'no' in Cell B162 and then enter the appropriate O&M unit costs ($ per kWh) in Row 164 (for the energy recovery systems chosen).

6.4.14 Energy Cost

The savings that can be generated using energy recovery technologies operating on biogas have been assumed to be due to energy savings. Therefore, the energy costs are handled separately from the O&M costs in this section.

6.4.14.1 Rows 173-179: Electricity Costs for Anaerobic Digester Operation

The electricity required for anaerobic digester operation is either that estimated for mixing by the spreadsheet or that which you entered earlier in Cell B69. Only Row 177 requires an entry from you. In Cell B177, enter the percentage of normal-priced electricity (versus peak priced electricity). For simplicity, we recommend entering the same value as you entered in Cell B122 above. Do not modify any other rows. The annual cost of electricity to operate just the anaerobic digesters is shown in Row 179.

6.4.14.2 Rows 181-190: Natural Gas Costs

If the value in Cell B182 is 0, you have offset all of your natural gas purchases with biogas and you should skip to Row 190 to make sure that Cell B190 is $0. Note that you may have chosen to purchase natural gas to augment your cogeneration capability (in Cell B243 of the Technical worksheet). If you have done that, the value in Cell B182 will not be zero.

If the value in Cell B182 is greater than 0, you must enter in Cell B182 the number of days per year that you will be purchasing extra natural gas. This feature has been added to the
spreadsheet to facilitate the difference in winter and summer heating demands. For example, if you determined that extra natural gas was needed only for winter heating (see discussion for Rows 154-159 of the Technical worksheet), then you could enter your winter average ambient temperature in Cell B157 of the Technical worksheet and then enter in Cell B182 of the Economics worksheet the number of days to be considered for winter. Alternatively, you may simply enter 365 for the entire year.

If the value in Cell B182 is greater than 0, you must also provide entries in Cells C187 and B188. These two cells allow for two prices of natural gas based on the volume purchased. If the natural gas price does not change as a function of volume purchased, simply enter into Cell C187 whatever price is showing in Cell B187 and then enter 0 in Cell B188.

**Do not modify any other rows.** The annual cost of natural gas purchased is shown in Row 190.

### 6.4.14.3 Rows 192-197: Diesel Costs

Enter in Cell B193 the amount of diesel (liters per year) used onsite for digester heating. If this number is not 0, we recommend entering 'no' in Cell B194 to decline our default diesel price and then entering the diesel price ($ per liter) in Cell C195. If you enter 0 in Cell B193, we recommend entering 'yes' into Cell B194. The annual cost of diesel purchased is shown in Row 197.

### 6.4.14.4 Row 199: Total Annual Energy Costs

**Do not modify this row.** Cell B199 shows the total annual energy costs for anaerobic digestion and energy recovery system-related energy uses only. The energy costs for the entire facility are not estimated because they do not affect the economic evaluation for the technologies of interest.

### 6.4.15 Economic Evaluation Measures

The remaining rows in the Economics worksheet (Rows 201-242) provide results of the economic calculations being conducted by the spreadsheet. There are still several rows requiring input from you, the rest present results.

### 6.4.15.1 Row 202: Exclude Anaerobic Digestion Costs

The anaerobic digestion costs (capital and operating) can overwhelm the capital and operating costs of the energy recovery and gas pretreatment systems. If you are not changing aspects of the anaerobic digestion process, you should enter 'yes' in Cell B202. This will exclude the anaerobic digestion costs from the subsequent analyses, providing economic evaluations only for the energy recovery and gas pretreatment systems.
If you are interested in comparing the impact of different anaerobic digestion processes on the cost effectiveness of adding energy recovery systems, then you should enter 'no' in Cell B202. Interpretation of the subsequent results will be explained below.

6.4.15.2 Row 203
Do not modify this row. It is needed to implement your answer in Cell B202.

6.4.15.3 Rows 206-207: Total Capital and Replacement Costs
Do not modify these rows. Cell B206 shows the total initial capital costs for the technologies you have selected. Cell B207 shows the sum of the initial capital costs plus the replacement capital costs brought back to today's dollars.

6.4.15.4 Row 208: Initial Grant for Capital
You may receive grants from outside sources to help purchase, in particular, energy recovery systems to produce renewable electricity and reduce greenhouse gas emissions. Enter in Cell B208 the amount of the grants that you expect to receive.

Note: Do not neglect the opportunity to receive outside funding to support your energy recovery initiative. A grant equal to even 20-25% of the initial capital cost can make an energy recovery project economically viable.

6.4.15.5 Rows 209-210: Net Capital and Replacement Costs
Do not modify these rows. The entries in Cells B209 and B210 show the net capital costs and the net capital plus replacement costs (in today's dollars) after accounting for any grants received.

6.4.15.6 Row 212
Do not modify this row. This row calculates the factor needed to convert annual savings or costs into a present value. It depends on both the discount rate you entered in Cell B43 and the planning period you entered in Cell B80.

6.4.15.7 Rows 215-225: Operating Costs, Savings and Revenues
Do not modify these rows. These rows present the total annual operating costs, savings and revenues ($ per year) and the life cycle costs, savings and revenues (today's dollars). The life cycle present values depend on the result of Row 212 and will be used for determining the overall life cycle cost.

6.4.15.8 Rows 227-229: Payback Periods
Do not modify these rows. The simple payback period (years) for the initial capital is shown in Cell B228. This commonly used economic parameter is calculated as the initial capital cost divided the savings and revenues minus the operating costs.
The simple payback period (years) including the replacement capital required over the planning period of the project is shown in Cell B229. This parameter is harder to interpret than the payback for the initial capital only. From a practical standpoint, if the payback period for the initial capital is shorter than the service life of the equipment, the result in Cell B229 may be ignored.

6.4.15.9 Row 231: Total Life Cycle Cost

The result in Cell B231 provides the total life cycle cost in today's dollars for the project being examined. The planning period and discount rate must be specified to appropriately consider the time value of money.

If the total life cycle cost is positive, then the project costs more than it returns over the planning period of interest. From a practical standpoint, if you have included the capital and operating costs of anaerobic digestion, the life cycle cost will probably be positive. The energy savings and revenues from using energy recovery systems from biogas will rarely if ever cover the capital cost of constructing anaerobic digesters. Nevertheless, you may compare the life cycle costs between different options to determine which is more cost effective. For example, if you are considering upgrading from mesophilic to thermophilic treatment, the option with the lower life cycle cost would be considered more cost-effective. Note that the spreadsheet does not consider biosolids disposal costs/savings, it considers only the costs/savings associated with energy recovery systems. This means that what you are seeing in the comparison is whether the additional biogas produced by increasing the VSR with thermophilic treatment provides enough extra energy and savings to offset the increased energy requirements and possible capital and operating cost increases.

If the total life cycle cost is negative, then the project returns more than it costs over the planning period of interest. From a practical standpoint, implementation of energy recovery systems for biogas (assuming no changes in anaerobic digestion) is not cost-effective if the life cycle cost is positive. This will also be reflected in a long payback period. When comparing two alternatives with negative life cycle costs, although both alternatives are cost-effective, the alternative with the lower life cycle cost is a better economic choice.

6.4.15.10 Rows 233 and 235: Net Annualized Costs

Another economic parameter for evaluation is the net annualized cost. Row 233 reports the capital cost on an annualized basis for the planning period and discount rate selected earlier. The net total annualized cost shown in Cell B234 subtracts the annual savings and revenues and adds the annual operating costs to the annualized capital cost. Similarly to the total life cycle cost, the net total annualized cost will very probably be positive if anaerobic digestion costs are considered. In the absence of anaerobic digestion costs, the net total annualized cost should be less than or equal to 0 for the alternative being considered to be cost-effective.
6.4.15.11 Rows 237-242: Internal Rate of Return

The total life cycle cost uses the discount rate and the planning period entered earlier in the spreadsheet. The internal rate of return is the discount rate that just makes the life cycle cost equal to zero for a given time frame of interest.

The IRR can be estimated iteratively with the spreadsheet if the life cycle cost calculated in Cell B231 is negative. If the life cycle cost in Cell B231 is positive, the internal rate of return has no practical meaning. To estimate the IRR, enter a time frame of interest in Cell B238. This must be greater than the simple payback period. Then enter an estimated IRR in Cell B239. Compare the factor calculated in Cell B241 to the target value in Cell B242. Increase or decrease the IRR estimate in Cell B239 until the factor calculated in Cell B241 is approximately equal to the target value in Cell B242. At that point, the value in Cell B239 can be taken as the IRR for the project.

Note that the IRR calculated here is only an estimate. Because of limitations in the spreadsheet set-up, the IRR includes only initial capital, not replacement costs. Additionally, IRR can be a difficult parameter to appropriately compare between projects, depending on the size of the capital expenditures and time scales involved. For this spreadsheet, we believe that the initial capital payback period, the total life cycle cost and the net total annualized economic cost are better comparison parameters than the IRR.

6.5 Case Studies

6.5.1 Introduction

The following case studies have been conducted to provide an indication of the type of results that the spreadsheet can produce and how to use those results. Additionally, enough information is provided for each case study to provide you an opportunity to practice using the spreadsheet. Indeed, we highly recommend using these case studies as an opportunity to familiarize yourself with the spreadsheet. Work through this user's manual in order as you enter the required information and make the choices required. Your results will likely differ slightly from those presented below depending on how you decide to implement some of the assumptions noted, but if you select the same capacities for the energy recovery systems as are shown in the results, your results should differ only by round-off error. If they do not, be sure to check your entries carefully. While the possibility of a spreadsheet error exists, we tested spreadsheet by having researchers of varying levels of experience conduct the cases and then compared the results. This allowed us to fix errors while at the same time indicating that a systematic, patient approach is needed to ensure all the needed entries are appropriately completed.

6.5.2 Background for All Cases

For all cases, when not given specific information, select default values. Assume no thickening of primary sludge and a 4% total solids (TS) with 60% volatile in the primary sludge. Use a biological solids retention time (SRT) for the upstream biological treatment of 5 days and select an aeration volume to provide a hydraulic retention time of 6 hours. Select a mixed liquor
suspended solids (MLSS) concentration equal to the estimate provided by the spreadsheet and use 80% as the %VS in the MLSS. Use a thickened biological sludge concentration of 4% and accept the default for flow. Choose the number of digesters so that the total volume of each digester is approximately 4,000 cubic meters. Use 65% methane in the raw biogas, hydrogen sulfide concentration of 300 ppm and siloxane concentration of 200 ppm. Assume the heat exchanger efficiency is 80%.

Do not send biogas offsite. Assume that hydrogen sulfide and siloxanes will be removed with an 85% removal rate in all cases, but carbon dioxide will not be removed in any cases. Use Iron Sponge for hydrogen sulfide removal and Activated carbon for siloxane removal. Use a discount rate of 5%, a service life for GPS of 20 years, and a planning period for the project of 20 years.

Assume that no excess heat is sold off-site. Do not consider salvage values as part of the revenue. Assume that the conversion efficiency of electricity into mechanical energy is 85% to account for the improved energy utilization efficiency of direct drives. If extra natural gas must be purchased, assume it will be purchased 365 days a year. Assume that purchased natural gas is always the price of interest (in other words, it does not change with the volume purchased). If peak electricity pricing is not mentioned, assume that 100% of the price is normal. Assume that no diesel is used at the facility. Assume that there is no grant for initial capital costs.

### 6.5.3 Comparing Energy Recovery Systems

A hypothetical plant is treating 20 million gallons a day in Illinois. It uses two-stage high rate, conventionally shaped mesophilic anaerobic digestion with a 10 day SRT in stage 1 and a 5-day SRT in stage two. Both stages are well mixed. The natural gas price is $0.25 per cubic meter and the electricity price is $0.09 per kWh.

Five different energy recovery options are considered. In one case, only boilers are used. The boiler capacity and number are selected to meet or slightly exceed the thermal energy needs of the digesters. If there is not enough biogas to meet that thermal energy need, extra natural gas is purchased until the boiler capacity is met. If excess biogas exists after meeting the thermal energy need, it is assumed to be flared. In the other four cases, engine cogeneration, microturbines, phosphoric acid (PA) fuels cells and direct drive technologies are used in conjunction with boilers. In these cases the energy recovery capacity is chosen to achieve 50% of the maximum capacity that could be achieved for that system. Boiler capacity is then added to meet thermal energy needs, if not met, and extra natural gas is purchased if needed to meet the total gas requirement from the combined electricity generation and boiler system.

Table 6-3 presents selected results from this comparison. Because only energy recovery options were considered, the capital and operating costs of the anaerobic digester were excluded. The spreadsheet estimates the capital costs of the required anaerobic digesters at over $41
million dollars. Clearly, the cost to construct and operate anaerobic digesters will not be readily recovered solely by adding energy recovery processes.

Table 6-3. Energy Recovery Options for 2-Stage Mesophilic Digestion.

<table>
<thead>
<tr>
<th>Predicted Parameter</th>
<th>Boiler</th>
<th>Engine Cogeneration</th>
<th>Microturbine</th>
<th>PA Fuel Cell</th>
<th>Direct Drive</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler Capacity (bhp)</td>
<td>125</td>
<td>70</td>
<td>55</td>
<td>50</td>
<td>60</td>
</tr>
<tr>
<td>Electricity Capacity (kW)</td>
<td>NAa</td>
<td>500</td>
<td>450</td>
<td>650</td>
<td>550</td>
</tr>
<tr>
<td>Biogas wasted (m³/day)</td>
<td>6,350</td>
<td>3,230</td>
<td>3,350</td>
<td>3,735</td>
<td>3,480</td>
</tr>
<tr>
<td>Fraction of Electricityb</td>
<td>NA</td>
<td>40%</td>
<td>36%</td>
<td>52%</td>
<td>52%</td>
</tr>
<tr>
<td>Net Capital Cost ($1,000)</td>
<td>205</td>
<td>1,418</td>
<td>1,909</td>
<td>5,749</td>
<td>1,761</td>
</tr>
<tr>
<td>O&amp;M ($1,000/yr)</td>
<td>100</td>
<td>274</td>
<td>258</td>
<td>364</td>
<td>565</td>
</tr>
<tr>
<td>Savings ($1,000/yr)</td>
<td>243</td>
<td>637</td>
<td>599</td>
<td>756</td>
<td>754</td>
</tr>
<tr>
<td>Payback time (yr)</td>
<td>1.4</td>
<td>3.9</td>
<td>5.6</td>
<td>14.7</td>
<td>9.4</td>
</tr>
</tbody>
</table>

aNA=Not applicable
bFraction of electricity produced and used onsite or fraction of electricity offset in the case of direct drives.

The savings shown in Table 6-3 for the Boiler are due to natural gas that no longer needs to be purchased. Clearly, boilers are cost-effective from a payback time standpoint. The payback times for the other options in this comparison include the savings due to both natural gas for heating and electricity. A different analysis would be required to find the payback to just add engine cogeneration, for example, if boilers were already on site. The life cycle cost results of the spreadsheet may be helpful for that analysis. Regardless, at $0.09 per kWh, the payback times for the other energy recovery alternatives are substantially longer than for the boiler alone. The PA fuel cell has a particularly long payback time because of its much higher capital cost. The payback time for the direct drive is longer than might be expected but this is due to the default operating and maintenance costs being substantially higher. This clearly indicates the importance of finding the best estimates possible for the terms in the model. Our default values are useful for preliminary analysis, but your site-specific values based on vendor quotes should be used for more detailed analysis.

In all cases, excess biogas was produced because of the constraint in the set-up that energy recovery technologies would be limited to 50% of the predicted maximum capacity. The analysis to determine the proper capacity to apply at a given site will require consideration of both summer and winter heating needs. The following cases consider some of the other factors that can play a role in determining capacity.

6.5.4 Converting to Thermophilic Digestion

The same hypothetical plant considers switching from two-stage mesophilic to one-stage thermophilic digestion. To retain approximately the same VS Reduction (%VSR), the thermophilic SRT is taken as 10 days. The same energy recovery alternatives are considered. Table 6-4 presents selected results from this comparison.
Table 6-4. Energy Recovery Options for 1-Stage Thermophilic Digestion.

<table>
<thead>
<tr>
<th>Predicted Parameter</th>
<th>Boiler</th>
<th>Engine Cogeneration</th>
<th>Microturbine</th>
<th>PA Fuel Cell</th>
<th>Direct Drive</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler Capacity (bhp)</td>
<td>250</td>
<td>190</td>
<td>180</td>
<td>170</td>
<td>190</td>
</tr>
<tr>
<td>Electricity Capacity (kW)</td>
<td>NAa</td>
<td>500</td>
<td>450</td>
<td>650</td>
<td>550</td>
</tr>
<tr>
<td>Biogas wasted (m³/day)</td>
<td>835</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Natural gas req’d (m³/day)</td>
<td>0</td>
<td>1,340</td>
<td>1,410</td>
<td>1,010</td>
<td>1,180</td>
</tr>
<tr>
<td>Fraction of Electricityb</td>
<td>NA</td>
<td>40%</td>
<td>36%</td>
<td>52%</td>
<td>52%</td>
</tr>
<tr>
<td>Net Capital Cost ($1,000)</td>
<td>410</td>
<td>1,615</td>
<td>2,113</td>
<td>5,945</td>
<td>1,958</td>
</tr>
<tr>
<td>O&amp;M ($1,000/yr)</td>
<td>94</td>
<td>389</td>
<td>380</td>
<td>449</td>
<td>666</td>
</tr>
<tr>
<td>Savings ($1,000/yr)</td>
<td>489</td>
<td>883</td>
<td>844</td>
<td>1,002</td>
<td>999</td>
</tr>
<tr>
<td>Payback time (yr)</td>
<td>1.0</td>
<td>3.3</td>
<td>4.6</td>
<td>10.8</td>
<td>5.9</td>
</tr>
</tbody>
</table>

a NA=Not applicable  
b Fraction of electricity produced and used onsite or fraction of electricity offset in the case of direct drives.

Only the boiler can meet the increased thermal energy needs without purchasing additional natural gas. The payback times of the other options are shorter than for the mesophilic case, but this is due to the net increased savings due to offsetting natural gas to provide all of the thermal energy needs. The relative ranking of the technologies in terms of payback time remains the same as for the mesophilic case.

Although not shown in Table 6-4, the capital cost of the digester construction is predicted by the spreadsheet to be $28.6 million dollars, slightly higher than two-thirds of the cost to construct the mesophilic digesters. This savings would arise from the decreased volume required to achieve the shorter SRT. If the facility was switching from two-stage mesophilic to 1-stage thermophilic and wanted to keep the same SRT, the %VSR would increase, from 56% to 61% for the conditions specified for this case. This would provide more biogas to meet both heating and electricity generation needs.

6.5.5 Considering Advanced Sludge Treatment

Instead of switching to thermophilic treatment, the hypothetical plant decides to consider an advanced sludge treatment add-on such as sonification in conjunction with two-stage mesophilic digestion. The add-on is assumed to augment digestion sufficient to achieve a %VSR of 85%.

Selected results are presented in Table 6-5 for the same five energy recovery options. These results must be interpreted carefully because the spreadsheet does not automatically add the capital and operating costs for the add-on. These can be considered by, for example, replacing the anaerobic digester cost options with values for the add-on and then choosing not to exclude the anaerobic digester capital and operating costs. If this is done, the capital costs of the
anaerobic digester will not be determined, but that is easily remedied by using a separate spreadsheet.

<table>
<thead>
<tr>
<th>Predicted Parameter</th>
<th>Boiler</th>
<th>Engine Cogeneration</th>
<th>Microturbine</th>
<th>PA Fuel Cell</th>
<th>Direct Drive</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler Capacity (bhp)</td>
<td>125</td>
<td>35</td>
<td>25</td>
<td>10</td>
<td>25</td>
</tr>
<tr>
<td>Electricity Capacity (kW)</td>
<td>NA&lt;sup&gt;a&lt;/sup&gt;</td>
<td>800</td>
<td>650</td>
<td>1,000</td>
<td>850</td>
</tr>
<tr>
<td>Biogas wasted (m³/day)</td>
<td>12,608</td>
<td>7,705</td>
<td>8,220</td>
<td>8,560</td>
<td>8,140</td>
</tr>
<tr>
<td>Fraction of Electricity&lt;sup&gt;b&lt;/sup&gt;</td>
<td>NA</td>
<td>63%</td>
<td>52%</td>
<td>79%</td>
<td>79%</td>
</tr>
<tr>
<td>Net Capital Cost ($1,000)</td>
<td>205</td>
<td>2,143</td>
<td>2,668</td>
<td>8,735</td>
<td>2,611</td>
</tr>
<tr>
<td>O&amp;M ($1,000/yr)</td>
<td>100</td>
<td>378</td>
<td>328</td>
<td>506</td>
<td>819</td>
</tr>
<tr>
<td>Savings ($1,000/yr)</td>
<td>243</td>
<td>873</td>
<td>756</td>
<td>1,032</td>
<td>1,032</td>
</tr>
<tr>
<td>Payback time (yr)</td>
<td>1.4</td>
<td>4.3</td>
<td>6.2</td>
<td>16.6</td>
<td>12.3</td>
</tr>
</tbody>
</table>

<sup>a</sup>NA=Not applicable  
<sup>b</sup>Fraction of electricity produced and used onsite or fraction of electricity offset in the case of direct drives.

In this case, the artificial limitation placed on the cases to use only 50% of the predicted maximum capacity may not be appropriate. For example, if the engine cogeneration capacity is increased to 1,500 kW, there would still be excess biogas produced (approximately 1,400 m³/day) and all of the facilities onsite electricity needs would be met. At $0.09 per kWh, however, the electricity price is too low for this to be cost-effective because the payback time for 1,500 kW increases to 7.3 years (the savings due to natural gas remained the same but the savings due to electricity did not increase proportionally to the increased capital cost of the engines).

The actual savings expected, however, should be significantly higher if the reduction in biosolids disposal costs arising from increased %VSR is considered. Because of the highly site-specific nature of biosolids disposal costs and the wide range of biosolids reuse and disposal options, the spreadsheet does not consider these and focuses entirely on the energy aspects.

### 6.5.6 Electricity Price Peaking

The hypothetical plant does not pay the same electricity price at all times. The same five energy recovery alternatives with two-stage mesophilic digestion are examined assuming the same normal electricity price of $0.09 per kWh but also a peak electricity price of $0.18 per kWh, occurring 20% of the time. The results are presented in Table 6-6.
Table 6-6. Impact of Electricity Peaking for 2-Stage Mesophilic Digestion.

<table>
<thead>
<tr>
<th>Predicted Parameter</th>
<th>Boiler</th>
<th>Engine Cogeneration</th>
<th>Microturbine</th>
<th>PA Fuel Cell</th>
<th>Direct Drive</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler Capacity (bhp)</td>
<td>125</td>
<td>70</td>
<td>55</td>
<td>50</td>
<td>60</td>
</tr>
<tr>
<td>Electricity Capacity (kW)</td>
<td>NAa</td>
<td>500</td>
<td>450</td>
<td>650</td>
<td>550</td>
</tr>
<tr>
<td>Biogas wasted (m³/day)</td>
<td>6,350</td>
<td>3,230</td>
<td>3,350</td>
<td>3,735</td>
<td>3,480</td>
</tr>
<tr>
<td>Fraction of Electricityb</td>
<td>NA</td>
<td>40%</td>
<td>36%</td>
<td>52%</td>
<td>52%</td>
</tr>
<tr>
<td>Net Capital Cost ($1,000)</td>
<td>205</td>
<td>1,418</td>
<td>1,909</td>
<td>5,749</td>
<td>1,761</td>
</tr>
<tr>
<td>O&amp;M ($1,000/yr)</td>
<td>116</td>
<td>290</td>
<td>274</td>
<td>380</td>
<td>581</td>
</tr>
<tr>
<td>Savings ($1,000/yr)</td>
<td>243</td>
<td>716</td>
<td>669</td>
<td>858</td>
<td>856</td>
</tr>
<tr>
<td>Payback time (yr)</td>
<td>1.6</td>
<td>3.3</td>
<td>4.8</td>
<td>12.0</td>
<td>6.4</td>
</tr>
</tbody>
</table>

aNA=Not applicable

bFraction of electricity produced and used onsite or fraction of electricity offset in the case of direct drives.

The capital costs remain the same as in the original case (Table 6-5) but the savings increased significantly more than the costs, leading to decreased payback times for all of the electricity generating systems. Because boilers do not produce electricity, the payback time actually decreases because of the slight increase in costs. Not surprisingly, the systems that produce the biggest electricity savings, the PA fuel cell and direct drive, show the biggest decrease in payback times, although the relative order of cost-effectiveness remains the same. These results simulate those expected with increases in normal electricity prices as well, with the differences between the payback periods decreasing as the more expensive technologies with greater electricity savings show greater decreases in payback periods. Under the right combination of electricity prices and onsite electricity production, the direct drive, in particular, but also possibly the fuel cell (at very high electricity prices) may become more cost-effective than the other two electricity production systems considered in these cases.

6.5.7 Final Comment on Case Studies

The results presented here are only a tiny fraction of the possibilities that can be considered. The LCAMER spreadsheet also provides life cycle costing information and emissions estimates for the different options. Furthermore, a wide number of different combinations can be considered. If you have been successful in duplicating the results presented here, you should now be ready to begin using the spreadsheet to examine your specific needs.
CHAPTER 7.0

RECOMMENDATIONS

Based on the results achieved to date with the spreadsheet model, it is recommended that:

17. The anaerobic digestion technical functions be modified in future work to consider factors such as digester feeding regime in addition to SRT and digester shape.

18. Where anaerobic digester, gas pretreatment or energy recovery process performance data or costs are based on limited survey data, additional effort should be made to increase the database, and hence to refine the data used in the spreadsheet model.

19. The LCC framework for LCAMER be expanded to include other unit processes, including (a) solids processing prior to anaerobic digestion and energy recovery and (b) biosolids disposal because they are key factors in the overall biosolids management strategy, and will provide a better overall assessment of the LCC.

20. The LCC framework for LCAMER be further expanded to include energy consuming unit processes (such as pump, aeration, etc.) related to handling liquid fraction of the WWTP.

21. Additional survey data be collected and tested to validate the spreadsheet model. Causes of significant deviations between reported and predicted results should be investigated.

22. The LCC framework model be expanded to include other external costs such as social and health costs, so that the model can better reflect the true LCC of a given scenario.

23. Because the costing data in the returned surveys were among the weakest elements of the study, facilities be encouraged to improve the quality of plant and process cost data.
# APPENDIX A

## SURVEY QUESTIONNAIRE TEMPLATE

- **Water Environment Research Foundation Research Study**

  "Cost Effective Energy Recovery from Anaerobically Digested Wastewater Biosolids"

  **NOTE**: Please try to provide data for 2002 (if not available, please state the year used)

<table>
<thead>
<tr>
<th>Background Information</th>
</tr>
</thead>
<tbody>
<tr>
<td>Name:</td>
</tr>
<tr>
<td>Today's Date:</td>
</tr>
<tr>
<td>Plant Name:</td>
</tr>
<tr>
<td>Plant Location:</td>
</tr>
<tr>
<td>Total Number of Employees</td>
</tr>
<tr>
<td>Total Number of Administration Employees</td>
</tr>
<tr>
<td>Total Number of Maintenance Employees</td>
</tr>
<tr>
<td>Population Serviced</td>
</tr>
<tr>
<td>Separate Stormwater Collection System or Combined with Wastewater?</td>
</tr>
<tr>
<td>Is the Plant in an Ozone Non-Attainment Area?</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Process Information</th>
</tr>
</thead>
<tbody>
<tr>
<td>Process Type (e.g. conventional activated sludge, trickling filter, etc.)</td>
</tr>
<tr>
<td>Sludge Treatment Type (e.g. anaerobic digestion, heat treatment, etc.)</td>
</tr>
<tr>
<td>Sludge Disposal Method</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Purchased Energy</th>
</tr>
</thead>
<tbody>
<tr>
<td>Annual Electricity Consumption (kWh/yr)</td>
</tr>
<tr>
<td>Price of Electricity ($ per kWh or annual cost)</td>
</tr>
<tr>
<td>Annual Natural Gas Consumption (1000 cu. ft/yr)</td>
</tr>
<tr>
<td>Price of Natural Gas ($ per 1000 cu.ft/yr or annual cost)</td>
</tr>
<tr>
<td>Other Fuel (e.g. Diesel) Annual Consumption (gal/yr)</td>
</tr>
<tr>
<td>Price of Other Fuel ($ per gal. or annual cost)</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Process Measurements</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mean Annual Influent BOD5 Concentration (mg/L)</td>
</tr>
<tr>
<td>Mean Annual Primary Effluent BOD5 Concentration (mg/L)</td>
</tr>
<tr>
<td>Mean Annual Secondary Effluent BOD5 Concentration (mg/L)</td>
</tr>
<tr>
<td>Mean Annual Influent TSS Concentration (mg/L)</td>
</tr>
<tr>
<td>Mean Annual Primary Effluent TSS Concentration (mg/L)</td>
</tr>
<tr>
<td>Mean Annual Secondary Effluent TSS Concentration (mg/L)</td>
</tr>
</tbody>
</table>

| Aeration Process Details |

---

*Cost-effective Energy Recovery from Anaerobically Digested Wastewater Solids*
<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>30</td>
<td>Average Daily Flow (mgd)</td>
</tr>
<tr>
<td>33</td>
<td>Plant Capacity (mgd)</td>
</tr>
<tr>
<td>34</td>
<td>Aeration Basin Volume (gal)</td>
</tr>
<tr>
<td>36</td>
<td>MLSS Concentration (mg/L)</td>
</tr>
<tr>
<td>37</td>
<td>MLVSS Concentration (mg/L)</td>
</tr>
<tr>
<td>38</td>
<td>RAS TSS Concentration (mg TSS/L)</td>
</tr>
<tr>
<td>39</td>
<td>RAS Flow Rate (gpd)</td>
</tr>
<tr>
<td>40</td>
<td>Aeration F:M Ratio</td>
</tr>
<tr>
<td>41</td>
<td>Solids Retention Time, SRT (d)</td>
</tr>
<tr>
<td>42</td>
<td><strong>Thickening Details</strong></td>
</tr>
<tr>
<td>43</td>
<td>Sludge Flow Rate from Primary Clarifier (gpd)</td>
</tr>
<tr>
<td>44</td>
<td>% Total Solids in Primary Sludge</td>
</tr>
<tr>
<td>45</td>
<td>% Volatile Solids in Primary Sludge</td>
</tr>
<tr>
<td>47</td>
<td>Primary Sludge Thickening Method (none, gravity, etc.)</td>
</tr>
<tr>
<td>48</td>
<td>Sludge Flow Rate from Primary Thickener (gpd)</td>
</tr>
<tr>
<td>49</td>
<td>% Total Solids in Thickened Primary Sludge</td>
</tr>
<tr>
<td>50</td>
<td>% Volatile Solids in Thickened Primary Sludge</td>
</tr>
<tr>
<td>51</td>
<td>WAS Flow Rate from Secondary Clarifier (gpd)</td>
</tr>
<tr>
<td>55</td>
<td>WAS Thickening Method (none, DAF, centrifuge, etc.)</td>
</tr>
<tr>
<td>56</td>
<td>Sludge Flow Rate from WAS Thickener (gpd)</td>
</tr>
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<td>57</td>
<td>% Total Solids in Thickened WAS</td>
</tr>
<tr>
<td>58</td>
<td>% Volatile Solids in Thickened WAS</td>
</tr>
<tr>
<td>59</td>
<td><strong>Anaerobic Digestion Process Details</strong></td>
</tr>
<tr>
<td>60</td>
<td>Anaerobic Process Type (please check one)</td>
</tr>
<tr>
<td>61</td>
<td>___ High-rate mesophilic</td>
</tr>
<tr>
<td>62</td>
<td>___ High-rate Thermophilic</td>
</tr>
<tr>
<td>63</td>
<td>___ Temperature-phased</td>
</tr>
<tr>
<td>64</td>
<td>___ Two-phase (Acid-gas phased)</td>
</tr>
<tr>
<td>65</td>
<td>___ Other (please specify): ______________________</td>
</tr>
<tr>
<td>65a</td>
<td>Feeding Regime (continuous or intermittent)</td>
</tr>
<tr>
<td>67</td>
<td>Primary Digesters</td>
</tr>
<tr>
<td>66</td>
<td>Process Operation Temperature (F)</td>
</tr>
<tr>
<td>66a</td>
<td>Digester pH</td>
</tr>
<tr>
<td>68</td>
<td>Mean Digester HRT (d)</td>
</tr>
<tr>
<td>69</td>
<td>Number of Digesters</td>
</tr>
<tr>
<td>70</td>
<td>Number of Digesters in Operation</td>
</tr>
<tr>
<td>71</td>
<td>Digester Shape (conventional, egg, etc.)</td>
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<tr>
<td>72</td>
<td>Digester Unit Volume (gal)</td>
</tr>
<tr>
<td>73</td>
<td>Digester Unit Working Volume (gal)</td>
</tr>
<tr>
<td>74</td>
<td>Digester Dimensions</td>
</tr>
<tr>
<td>89</td>
<td>% Total Solids in Digested Sludge</td>
</tr>
<tr>
<td>90</td>
<td>% Volatile Solids in Digested Sludge</td>
</tr>
<tr>
<td>91</td>
<td>% Volatile Solids Reduction</td>
</tr>
<tr>
<td>75</td>
<td>Type of Mixing System (gas, pump, etc.)</td>
</tr>
<tr>
<td>76</td>
<td>Cover Type (fixed, floating, etc.)</td>
</tr>
<tr>
<td>77</td>
<td>Cover Material (e.g. steel)</td>
</tr>
<tr>
<td>80</td>
<td>Wall Material (e.g. concrete, steel)</td>
</tr>
<tr>
<td>83</td>
<td>Wall Insulation Thickness (inches)</td>
</tr>
<tr>
<td>84</td>
<td>Total Number of Heat Exchangers</td>
</tr>
<tr>
<td>85</td>
<td>Heat Exchanger Type (e.g. spiral-plate, tube-in-tube, water bath, steam inject)</td>
</tr>
<tr>
<td>86</td>
<td>Primary Digesters Go Off-line every ___ Year(s) for Maintenance Work</td>
</tr>
<tr>
<td>87</td>
<td>Secondary Digesters Go Off-line every ___ Year(s) for Maintenance Work</td>
</tr>
<tr>
<td>88</td>
<td>Heat Exchangers Go Off-line every ___ Year(s) for Maintenance Work</td>
</tr>
<tr>
<td>92</td>
<td>Anaerobic Digester Gas Production (cu. ft/d)</td>
</tr>
<tr>
<td>93</td>
<td>Anaerobic Digester Gas % Methane</td>
</tr>
<tr>
<td>94</td>
<td>Anaerobic Digester Gas % CO₂</td>
</tr>
<tr>
<td>95</td>
<td>Anaerobic Digester Gas % H₂O</td>
</tr>
<tr>
<td>96</td>
<td>Anaerobic Digester Gas ppm H₂S</td>
</tr>
<tr>
<td>97</td>
<td>Anaerobic Digester Gas % Other (please specify)</td>
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<tr>
<td>98</td>
<td>Volume of Digester Gas Lost due to System Leakage (cu. ft/d)</td>
</tr>
<tr>
<td>99</td>
<td>Volume of Digester Gas Flared (cu. ft/d)</td>
</tr>
<tr>
<td>100</td>
<td>Volume of Digester Gas Used for Energy (cu. ft/d)</td>
</tr>
<tr>
<td>102</td>
<td>Mass of Solids Removed from Plant (ton/d)</td>
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<tr>
<td>103</td>
<td>Gas Pretreatment/Scrubbing Details</td>
</tr>
<tr>
<td>104</td>
<td>Crude Digester Gas Flow Rate (cu. ft/d)</td>
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<tr>
<td>106</td>
<td>Purified Digester Gas Flow Rate (cu. ft/d)</td>
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<tr>
<td>118</td>
<td>Average Purified Digester Gas % Methane</td>
</tr>
<tr>
<td>119</td>
<td>Average Purified Digester Gas % CO₂</td>
</tr>
<tr>
<td>120</td>
<td>Average Purified Digester Gas % H₂O</td>
</tr>
<tr>
<td>121</td>
<td>Average Purified Digester Gas ppm H₂S</td>
</tr>
<tr>
<td>122</td>
<td>Average Purified Digester Gas % Others (please specify)</td>
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<td>108</td>
<td>H₂S Removal Process (if applicable)</td>
</tr>
<tr>
<td>109</td>
<td>Chemical Dosage for H₂S Removal (please specify, lb/d)</td>
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<td>110</td>
<td>H₂S Removal Equipment Goes Off-line Every ___ Year(s) for Maintenance</td>
</tr>
<tr>
<td>111</td>
<td>Frequency of H₂S Removal Equipment Regeneration</td>
</tr>
<tr>
<td>112</td>
<td>Regenerating Chemical(s) Dosage Per Regeneration (please specify, lb)</td>
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<td>Page</td>
<td>Description</td>
</tr>
<tr>
<td>------</td>
<td>-------------</td>
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<tr>
<td>113</td>
<td>CO₂ Removal Process (if applicable)</td>
</tr>
<tr>
<td>114</td>
<td>Chemical Dosage for CO₂ Removal (please specify, lb/d)</td>
</tr>
<tr>
<td>115</td>
<td>CO₂ Removal Equipment Goes Off-line Every ___ Year(s) for Maintenance</td>
</tr>
<tr>
<td>116</td>
<td>Frequency of CO₂ Equipment Regeneration</td>
</tr>
<tr>
<td>117</td>
<td>Regenerating Chemical(s) Dosage Per Regeneration (please specify, lb)</td>
</tr>
<tr>
<td>123</td>
<td><strong>Energy Recovery Process Details</strong></td>
</tr>
<tr>
<td>124</td>
<td>Energy Recovery Process (please check one)</td>
</tr>
<tr>
<td>125</td>
<td>___ None (biogas is flared)</td>
</tr>
<tr>
<td>126</td>
<td>___ Boilers (heat recovery)</td>
</tr>
<tr>
<td>127</td>
<td>Please Specify the Type, Number and Size of Boilers Being Used:</td>
</tr>
<tr>
<td>128</td>
<td>___ Gas engine cogeneration: naturally aspirated, spark-ignited</td>
</tr>
<tr>
<td>129</td>
<td>Please Specify the Type, Number and Size of Gas Engine Being Used:</td>
</tr>
<tr>
<td>130</td>
<td>___ Direct Drive Engines</td>
</tr>
<tr>
<td>131</td>
<td>Please Specify the Type, Number and Size of Direct Drives Being Used:</td>
</tr>
<tr>
<td>132</td>
<td>___ Microturbine(s) cogeneration</td>
</tr>
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<td>133</td>
<td>Please Specify the Type, Number and Size of Microturbine(s) Being Used:</td>
</tr>
<tr>
<td>134</td>
<td>___ Fuel cell(s)</td>
</tr>
<tr>
<td>135</td>
<td>Please Specify the Type, Number and Size of Fuel Cell(s) Being Used:</td>
</tr>
<tr>
<td>136</td>
<td>___ Other (please specify type and total number of unit):</td>
</tr>
<tr>
<td>137</td>
<td>Natural Gas Used in Summer Months (cu.ft/d)</td>
</tr>
<tr>
<td>138</td>
<td>Natural Gas Used in Winter Months (cu.ft/d)</td>
</tr>
<tr>
<td>139</td>
<td>Digester Gas Used in Summer Months (cu.ft/d)</td>
</tr>
<tr>
<td>140</td>
<td>Digester Gas Used in Winter Months (cu.ft/d)</td>
</tr>
<tr>
<td>141</td>
<td>Equipment Goes Off-line Every ___ Year(s) for Maintenance</td>
</tr>
<tr>
<td>142</td>
<td>Thermal Energy Recovered (MBTU/d)</td>
</tr>
<tr>
<td>143</td>
<td>Thermal Energy Efficiency (%)</td>
</tr>
<tr>
<td>144</td>
<td>Electrical Energy Recovered (kWh/d)</td>
</tr>
<tr>
<td>145</td>
<td>Electrical Energy Efficiency (%)</td>
</tr>
<tr>
<td>146</td>
<td>Total Design Capacity of Cogeneration Equipment (kW)</td>
</tr>
<tr>
<td>169</td>
<td><strong>Capital and Operating Costs</strong></td>
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<tr>
<td>170</td>
<td>Anaerobic Digester Capital Cost (US$)</td>
</tr>
<tr>
<td>171</td>
<td>Year of Purchase for Anaerobic Digesters</td>
</tr>
<tr>
<td>-----</td>
<td>----------------------------------------</td>
</tr>
<tr>
<td>174</td>
<td>H₂S Removal Equipment Capital Cost (US$)</td>
</tr>
<tr>
<td>175</td>
<td>CO₂ Removal Equipment Capital Cost (US$)</td>
</tr>
<tr>
<td>176</td>
<td>Year of Purchase for Gas Pretreatment Equipment</td>
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<tr>
<td>177</td>
<td>Boilers Capital Cost (US$)</td>
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<tr>
<td>178</td>
<td>Year of Purchase for Boilers</td>
</tr>
<tr>
<td>179</td>
<td>Cogeneration/Fuel Cell Equipment Capital Cost (US$)</td>
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<tr>
<td>180</td>
<td>Year of Purchase for Cogeneration Equipment/Fuel Cell</td>
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<td>181</td>
<td>Total Annual Operating Cost for Anaerobic Digesters (US$)</td>
</tr>
<tr>
<td></td>
<td>Labor Cost ____________   Maintenance Cost _____________   Chemical Cost ____________</td>
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<tr>
<td>182</td>
<td>Total Annual Operating Cost for H₂S Removal Equipment (US$)</td>
</tr>
<tr>
<td></td>
<td>Labor Cost ____________   Maintenance Cost _____________   Chemical Cost ____________</td>
</tr>
<tr>
<td>183</td>
<td>Total Annual Operating Cost for CO₂ Removal Equipment (US$)</td>
</tr>
<tr>
<td></td>
<td>Labor Cost ____________   Maintenance Cost _____________   Chemical Cost ____________</td>
</tr>
<tr>
<td>184</td>
<td>Total Annual Operating Cost for Boilers (US$)</td>
</tr>
<tr>
<td></td>
<td>Labor Cost ____________   Maintenance Cost _____________   Chemical Cost ____________</td>
</tr>
<tr>
<td>185</td>
<td>Total Annual Operating Cost for Cogeneration Equipment/Fuel Cell (US$)</td>
</tr>
<tr>
<td></td>
<td>Labor Cost ____________   Maintenance Cost _____________   Chemical Cost ____________</td>
</tr>
<tr>
<td>186</td>
<td>Annual Operating Savings due to Cogeneration/Fuel Cell (US$)</td>
</tr>
</tbody>
</table>
NOTES FROM PLANT SITE VISITS

Digestion

Port Dover mix their egg-shape digester with a jet pump, which they found is prone to clogging with hairballs, and need cleaning every six to 12 months. However, Terminal Island finds their egg-shaped digester performance satisfactory, with very little cleaning required and they also use pump for mixing.

Sludge-sludge heat exchangers had not worked at Annacis Island (who operates at thermophilic temperatures) and they switched to Sludge-to-hot water heat exchangers.

Gas Treatment

For the boilers at Port Dover, moisture traps is the only form of gas treatment employed, though ferrous chloride is added upstream.

West Point does not treat their gas for their direct drive and cogeneration engine units, other than using a chiller to remove water and oil. They have a 300ppm H2S limit for their cogeneration units, which they have no trouble meeting. The co-gen engines (1300kW each) were built in 1985 and are scheduled for replacement in 2007. They’ve had some maintenance problems due to the fact that they were low cost, high rpm units. The new engines will be lower rpm units with higher capital costs, but should have lower operating costs.

East Bay MUD only uses ferric chloride addition before digestion to reduce H2S. They use gas engines for cogeneration and do not mention any other form of gas treatment.

San Jose/Santa Clara only add FeCl2 to their headworks, and use not other form gas treatment for their gas engine cogeneration and direct drive units.

Annacis Island cools gas using a glycol system (from 55 to 8°C, from to 131 to 41°F) to remove water, then reheated to 50°C (122°F) and sent directly to boilers or to activated carbon units before cogeneration units. Previous to the carbon unit installation in 2001, there were extensive siloxane build-up problems. A disposal problem has arisen with disposal of the spent carbon; they were collecting arsenic. They had to change the carbon more often than required for siloxane removal, in order to meet the Canadian hazardous waste limit of 5ppm for arsenic disposal. But the limit has recently been lowered to 2.5ppm, so this is no longer an option.

Lewiston WPCP treats their biogas with a desiccant to remove water (~ $10, 000/yr) a rechargeable activated carbon unit (SAG™) ($1000/yr chemical cost), and about 0.38 m³/d (100 gal/day) of FeCl3 is added to the sludge to remove H2S. Previously, they had not treated their gas...
when used for gas engine co-generation, though they had to periodically replace damaged spark plugs (siloxanes were suspected to be the cause). Siloxane removal is required for the microturbines.

Columbia Blvd. treats gas with activated carbon to remove sulfur for their fuel cells, but do not treat gas going to microturbines, which may be the cause of scaling problems. Gas is sent using a low pressure compressor for the fuel cells and microturbine, and separate compressors send the digester gas into a pipeline for a company nearby, without treatment.

Renton currently scrubs their gas using Binax, add an odorant and highly compress the gas to send through a natural gas pipeline.

**Boilers**

Burlington Skyway WWTP recovers heat only, using boilers and they do not treat their biogas. About 10% of the tubing needed replacement over a course of 20 years. Natural gas is usually only used for boiler start up (the high number reported in the survey was due to a digester breakdown for 2 months). Heat recovered is used for digestion and other heating needs. In the summer, 70% is flared (50% in winter), and they would like to look into cogeneration. Labor associated maintenance involves checking the burners.

Port Dover uses boilers only for their biogas, and about 50% of the biogas is flared in the summer. Natural gas is used for start-up and supplementing the biogas. The mixture is about \( \frac{1}{2} \) and \( \frac{1}{2} \). Boilers do require an anti-rust agent. Port Dover uses Naschem boiler and steam liner treatment (~20L/year or 0.7063 scf/year, costs about $200). Burlington Skyway uses sodium nitrite and a rust inhibitor to pull out the free oxygen.

**Gas Engines**

Terminal Island is phasing out gas engines because they are failing to meet air emissions standards. To rectify by adding air pollution control systems would be too expensive. By end of 2004, digester will be used solely for boilers. There are plans to use purified digester gas in their fuel cells (which currently operate on natural gas only).

A timing problem with the gas engine units at Annacis Island had caused high temperatures and air emissions, but the problem had been adjusted.

**Microturbine**

The microturbine (1 unit) at Lewiston NY is being fed with biogas at 40kW and adding natural gas to get 60kW total. About 6,535 MJ (6.2 million BTU) of heat was recovered per day (the heat demand for digesters and digestion related buildings is 4,216-5,270 MJ (4-5 million BTU) in the summer and 6,324-8,432 MJ (6-8 million BTU) in the winter). If the gas production rate changes, the BTU change is sensed automatically compensates by changing the fuel/air ratio. Currently they are not running the microturbine at full capacity because economically, the additional natural gas purchase would not be worth it, given the current electricity prices. Part loading does not seem to affect the microturbine efficiency.
Energy Recovery – Savings/Revenues, and General Comments

Anacis Island has two sets of co-gen engines, the new set (Jenbacher) have computerized controls and are used preferentially. Older units are used if required. Heat recovered from engines is usually sufficient to supply all the hot water need. Boilers are rarely used. Electricity generated is used in plant only. In addition, their electrical rate is cheap because they have their own transformer.

At Lewiston WPCP the electricity rate is $0.06 + addition charges = $0.12 total/kW. They monitor their consumption carefully because demand charges are applied for the whole year when they exceed 150kW on any given day. They also have a 250kW diesel generator for peak shaving (can be cheaper when run for short periods).

All the plants visited in California are generally satisfied with their energy recovery scheme, which tends to be cost effective due to the high cost of electricity and natural gas there.

Oceanside uses gas engine cogeneration and considers their digester gas recovery “worthwhile and cost-effective”. Similar comments are made for East Bay MUD, Terminal Island,

Addition of microturbines at San Elijo did not require any staffing increase. In general they are satisfied with the system, which generates $40,000/yr of net savings due to the cogeneration.

The San Jose facility considers digester gas a valuable component in their energy conservation efforts. The savings through cogeneration alone is ~$6M/year.

At Chiquita, no supplemental natural gas is used because it is not available for the plant, as such this is considered the most cost effective alternative. The microturbines were initially NOT cost effective as they were fairly new technologies and required a series of modifications. However, they have since adapted to the technology, which has begun to generate some cost savings.

At Eastern MWD, 10% of digester gas is flared (after using 70% for blowers and 20% in microturbines). The plant may consider building a tank to store excess gas in the future to maximize gas usage.

Gas storage may be an important component of energy recovery. East Bay MUD uses low pressure gas holders and they only flare 1 to 2% of their digester gas, and recovering the remainder for energy. Furthermore, the plant has a policy not to burn NG, hence the power generation at the plant is 100% digester gas fuelled. Digester gas recovered provides 70-75% of the plants electricity needs.

Electricity generated at Columbia Blvd. is used on site only. They generate much less than they use. Electricity is generated using fuel cells and microturbines. Fuel cells can generate up to 1.4 million kWh/yr. Boilers are used as well as sending gas directly to a company nearby, still, %50 of the gas is flared.

West Point produced electricity that is fed directly into the grid (none is used at the plant). They have a system that offsets metering, based on total kWh in vs. out. The electricity they supplied was worth $800, 000 in 2003.
Renton South’s sale of their gas through the pipeline brings in revenue of $400,000 per year. A recent installation of fuel cells and future plants for turbines is expected to make the plant totally energy self-sufficient (with excess heat supplied to nearby industries).

Untreated gas sent by Columbia blvd. to a nearby company generates $100,000 per year. The fuel cells generate 1.4M kWh/yr, which values at $92000 as renewable power, or $58000 as “grid” supplied power.

**Funding**

Microturbine and fuel cell projects seem to have been heavily funded.

Eastern MWD obtained microturbines free through California Air Quality Improvement Rebate Program.

Chiquita received two out of four microturbines free from the state through air quality rebate. The remaining two were purchased for $1M total.

The microturbine unit at Lewiston WPCP is valued at $225,000, but was 100% funded through government grants.

San Elijo did not seem to receive funding assistance for their microturbines.

Fuel cells cost Renton South only $2 million out of $22 million due to grants from the U.S. EPA and the fuel cell manufacturer.

Fuel cells at Columbia Blvd were supported by the Oregon Office of Energy, Portland General Electric, the Fuel cell Climate Change Program and Oregon’s Business Energy Tax Credit Program, cutting the cost of $1.3M down to $790,000.

**References for Field Notes**


Christina Jacob, and Brian Hystad, Annacis Island Plant, Vancouver, BC. Notes from site visit by BAT Associates. March 17, 2004.


Dean Marriot, Director, Columbia Boulevard Wastewater Treatment plant – Brochure, August 2003.

Tim Lockhart, Chief Operator, Town of Lewiston Water Pollution Control Centre, Lewiston, NY. Site Visit October 30, 2003.

Ron Armstrong, Port Dover Wastewater Treatment Plant, Norfolk, ON, Site Visit October 2, 2003.


John Gibbs, Division Manager, San Jose/Santa Clara Water Pollution Control Plant. San Jose, CA. Notes from site visit by BAT Associates. April 23, 2004.
APPENDIX C

ASSESSMENT OF EFFECT OF CONVERSION TO ADVANCED ANAEROBIC DIGESTION PROCESSES FROM MESOPHILIC OPERATION

Literature Review

A comprehensive review of anaerobic digestion process upgrades in North America and Europe was prepared by Schafer et al. (2003). Most importantly for this assessment, these authors included details of the advanced digestion system operation, including comparisons of VS Reductions with baseline mesophilic operation, and with the newer advance digestion process.

Daigger (1998) presents the equations for ammonia and phosphate production by anaerobic treatment of wastewater solids. The equation for nitrogen release from organic solids is:

\[ C_5H_7O_2N + 4H_2O \rightarrow 1.5CO_2 + 2.5CH_4 + NH_4HCO_3 \]  
(C-1)

While that for phosphate release from the insoluble compound is

\[ MgKPO_4 \rightarrow Mg^{+2} + K^{+2} + PO_4^{-3} \]  
(C-2)

As an approximation, Daigger states that nitrogen release (as N) will be about 10% of the volatile suspended solids that are destroyed. He suggests that phosphorus is initially present at about 20% of the nitrogen concentration, so the phosphorus release will be about 1-2% of VSS destroyed. Complicating the phosphorus issue is precipitation with metals such as aluminum or iron that may be present in the solids from use as a phosphorus precipitation chemical in the liquid treatment train.

Based on the equations for nutrient conversion in anaerobic digestion provided by Daigger (1998), it is possible to estimate the effect of different anaerobic digester supernatant streams on liquid treatment performance, using mesophilic digestion as the baseline condition. The data provided by Schafer et al. (2003) allow for the assessment, using a number of initial assumptions, which in this assessment were:

Raw wastewater characteristics:

- BOD₅: 200 mg/L (0.0017 lb/gal)
- Total suspended solids: 200 mg/L (0.0017 lb/gal)
- Ammonia-N: 20 mg/L (0.0002 lb/gal)
Based on this set of conditions, it is possible to calculate the effect that digester supernatant might have on plant loadings if switching from conventional mesophilic operation to an advanced form of digestion, such as staged thermophilic, temperature-phased anaerobic digestion (TPAD), acid-gas phased digestion or a combination of the above. Given that conventional mesophilic digestion will contribute some additional loading of BOD₅, solids and nutrients, the question is to what extent the loadings will change in converting to an advanced form of anaerobic digestion. Tables C-1 through C-3 examine the increased loadings to the liquid train from the digestion process conversion, based on the difference in volatile solids reductions reported by Schafer et al. (2003).
<table>
<thead>
<tr>
<th>Plant</th>
<th>Mode</th>
<th>VS redn (%)</th>
<th>Supernatant Parameter</th>
<th>Conc’n mg/L (lb/gal)</th>
<th>% of Inf. Load</th>
<th>Load Change from Meso Operation</th>
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<td>Annacis Island, Vancouver, B.C.</td>
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<td>BOD₅</td>
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<td>NH₃-N</td>
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<td>TP</td>
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<td>Staged thermo</td>
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<td>BOD₅</td>
<td>1108 (0.0092)</td>
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<td>-27.6%</td>
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<td>NH₃-N</td>
<td>2087 (0.0174)</td>
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<td>TP</td>
<td>81.7 (0.007)</td>
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<td>Lion’s Gate</td>
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<td>57.6</td>
<td>BOD₅</td>
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<td>Vancouver, B.C.</td>
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<td>TSS</td>
<td>2045 (0.0171)</td>
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<td>NH₃-N</td>
<td>1926 (0.0161)</td>
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<td>TP</td>
<td>75.4 (0.006)</td>
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<td>Staged thermo</td>
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<td>BOD₅</td>
<td>960 (0.0080)</td>
<td>4.8</td>
<td>-21.3%</td>
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<td>1926 (0.0161)</td>
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<td>NH₃-N</td>
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<td>TP</td>
<td>81.7 (0.0007)</td>
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<td>TP</td>
<td>85.5 (0.0007)</td>
<td>12.2</td>
<td>5.2%</td>
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## Table C-3. Change in Contaminant Loadings to Liquid Processes due to Anaerobic Digestion Process Conversion to Acid-Gas-Phased Operation.

<table>
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<tr>
<th>Plant</th>
<th>Mode</th>
<th>VS Red’n (%)</th>
<th>Supernatant Parameter</th>
<th>Conc’n, mg/L (lb/gal)</th>
<th>% on Infl. Load</th>
<th>Load Change from Meso Operation</th>
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<tr>
<td>DuPage County, IL</td>
<td>Mesophilic</td>
<td>45</td>
<td>BOD₅</td>
<td>1560 (0.0130)</td>
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<td>BOD₅</td>
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<td>-29.5%</td>
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<td></td>
<td></td>
<td></td>
<td>TP</td>
<td>62.7 (0.0005)</td>
<td>9</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Acid-Gas T/M</td>
<td>50</td>
<td>BOD₅</td>
<td>1442 (0.0120)</td>
<td>7.2</td>
<td>-4.0%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TSS</td>
<td>2250 (0.0188)</td>
<td>11.3</td>
<td>-1.7%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>NH₃-N</td>
<td>1673 (0.0140)</td>
<td>83.6</td>
<td>4.5%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TP</td>
<td>65.5 (0.0005)</td>
<td>9.4</td>
<td>4.4%</td>
</tr>
</tbody>
</table>

*Cost-effective Energy Recovery from Anaerobically Digested Wastewater Solids*
<table>
<thead>
<tr>
<th>Plant</th>
<th>Mode</th>
<th>I1S Red’n (%</th>
<th>Parameter</th>
<th>Conc’n, mg/L (lb/gal)</th>
<th>% on Infl. Load</th>
<th>Load Change from Meso Operation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Inland Empire Util., CA</td>
<td>Mesophilic</td>
<td>54</td>
<td>BOD5</td>
<td>1335 (0.0111)</td>
<td>6.7</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TSS</td>
<td>2143 (0.0179)</td>
<td>10.7</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>NH3-N</td>
<td>1805 (0.0151)</td>
<td>90.3</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TP</td>
<td>70.7 (0.0006)</td>
<td>10.1</td>
<td></td>
</tr>
<tr>
<td>3-phase, A/T/M</td>
<td></td>
<td>56</td>
<td>BOD5</td>
<td>1286 (0.0107)</td>
<td>6.4</td>
<td>-4.5%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TSS</td>
<td>2093 (0.0175)</td>
<td>10.5</td>
<td>-1.9%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>NH3-N</td>
<td>1867 (0.0156)</td>
<td>93.4</td>
<td>3.4%</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>TP</td>
<td>73.1 (0.0006)</td>
<td>10.4</td>
<td>3.0%</td>
</tr>
</tbody>
</table>

T = thermophilic  
M = mesophilic  
TPAD = temperature-phased anaerobic digestion  
A = acid

This spreadsheet analysis confirms the observations of Daigger (1998) that ammonia-N is the contaminant in digester supernatant that may have the greatest impact on liquid train operation. Supernatant return, even with standard high-rate mesophilic anaerobic digestion, can contribute as much as 90% of the influent wastewater ammonia-N mass, but often in a shorter time period. Conversion of the digestion process from mesophilic to an advanced process has the potential to increase the ammonia-N and phosphorus loadings by 1% up to 50%, depending on the improvement in the volatile solids reduction with the new digestion process. Conversely, BOD5 and TSS loadings may be expected to decrease by 1% to 36% in returning supernatant of the advance processes relative to mesophilic operation, also due to the improved volatile solids reduction.

The actual effect on plant operation resulting from change in digester process supernatant cannot be determined based on the estimated change in BOD, TSS and nutrient loadings. For those plants that are operating at or near capacity, the change in digester process could have a significant adverse effect due to the increased ammonia-N loading. Plants with plenty of excess capacity may see only slight or no effect of switching digester processes. Simulation is required to determine if the change in loadings would actually lead to failure, as determined by falling pH or ammonia-N breakthrough. Such effects are best addressed using process modeling software.

**Process Modeling Assessment**

Efforts to assess the effect of supernatant from advanced anaerobic digestion processes on liquid train performance involved used of spreadsheets and simplified modeling expressions. To expand on this assessment, simulation was conducted with the software package BioWin32 version 1.2.1 at the University of Toronto.

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4 Mention of any commercial product or trade name does not reflect endorsement by the Research Team or WERF.

C-6
The characteristics of the influent wastewater used to produce the wastewater solids sent to the anaerobic digestion system were:

- flow 10,000 m$^3$/d (2.64 MGD)
- BOD = 200 mg/L (0.0017 lb/gal)
- TSS = 200 mg/L (0.0017 lb/gal)
- VSS = 75% of TSS
- TKN = 25 mg-N/L (0.0002 lb/gal)
- TP = 7 mg-P/L (0.0001 lb/gal)
- Alkalinity = 200 mg CaCO$_3$/L (0.0017 lb/gal)
- Mg = 30 mg/L (0.0003 lb/gal)

The initial results for nitrogen release calculated in the 2003 Annual Report (based on numbers from Daigger, 1998) were derived on the nitrogen content of biomass, formulated as C$_5$H$_7$O$_2$N. However, the BioWin32 simulator reveals that this procedure tends to overestimate the amount of nitrogen released when primary sludge (PS) is digested together with waste activated sludge (WAS). The specific nitrogen content (SNC) of any sludge can be determined by dividing the measured total Kjeldahl nitrogen by the volatile solids concentration. A summary of some SNC and specific phosphorus content (SPC) values is provided in Table C-4.

<table>
<thead>
<tr>
<th>Sludge</th>
<th>Specific Nitrogen Content g-N/g-VS (lb-N/lb-VS)</th>
<th>Specific Phosphorus Content g-P/g-VS (lb-P/lb-VS)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Based on cell composition C$_5$H$_7$O$<em>2$NP$</em>{0.09}$</td>
<td>0.1239 (0.1239)</td>
<td>0.0248 (0.0248)</td>
</tr>
<tr>
<td>Biowin simulated primary sludge</td>
<td>0.0403 (0.0403)</td>
<td>0.0203 (0.0203)</td>
</tr>
<tr>
<td>Biowin simulated WAS</td>
<td>0.1148 (0.1148)</td>
<td>0.0314 (0.0314)</td>
</tr>
</tbody>
</table>

The specific nitrogen content of WAS simulated by BioWin is very close to that based on theoretical cell composition. The BioWin simulated Primary Sludge has a much lower nitrogen content. This will have an effect when the estimated nitrogen released during anaerobic digestion is predicted. That is, the use of theoretical cell composition or WAS composition will overestimate the nitrogen released during anaerobic digestion when a significant portion of the sludge to be digested is primary sludge. The specific phosphorus content used in the simulation software is different for primary sludge and WAS, reflecting the difference in organic composition of the two waste streams. The specific phosphorus content based on cell composition lies near the average of the BioWin values for primary and waste activated sludge.

The nitrogen and phosphorus contained in both primary and waste activated sludges is practically all in the particulate form, as predicted by the software model. Because all the nitrogen
and phosphorus is in particulate form, an increase in the VS Reduction will increase the soluble N and P content proportionately to the sludge specific nitrogen and phosphorus contents. In BioWin, an increase in VS Reduction to a given VS Reduction representative of a particular advanced digestion technology is achieved by adjusting the heterotrophic anaerobic digestion hydrolysis rate.

The %VSR values in Table C-5 were used for the corresponding advanced digestion technologies.

<table>
<thead>
<tr>
<th>Digestion Technology</th>
<th>% VSR</th>
</tr>
</thead>
<tbody>
<tr>
<td>High rate mesophilic (single-stage)</td>
<td>47%</td>
</tr>
<tr>
<td>Thermophilic (single-stage)</td>
<td>63%</td>
</tr>
<tr>
<td>Temperature-phased (Thermo-Meso)</td>
<td>61%</td>
</tr>
<tr>
<td>Two-phase (Acid-Gase, both mesophilic)</td>
<td>58%</td>
</tr>
</tbody>
</table>

The digester supernatant recycle NH$_3$-N content was increased and the treatment plant effluent ammonia was simulated, with the results shown in Figure C-1. The simulation used a SRT of 10 days and a temperature of 15°C. The nitrification of the plant was determined to be alkalinity limited, and therefore alkalinity was added after the Bioreactor model. The amount of NH$_3$ recycled in supernatant was quantified as a percent of the total load to the plant. Figure C-2 shows that a higher contribution of recycled supernatant ammonia will increase the final effluent ammonia concentration to a greater extent when the nitrification capacity is diminished. When nitrification is proceeding well (e.g. as represented by the “10 kg/d (22 lb/d) alkalinity added” line in Figure C-2), the recycled ammonia can contribute from 10-15% of the total plant load without much effect. Beyond this level, or when nitrification is poor (e.g. as represented by the “0 kg/d alkalinity added” line in Figure C-2), strategies to improve nitrification should be investigated and/or separate treatment of the recycle stream before introduction to the liquid train should be investigated to prevent higher final effluent concentrations of ammonia.

![Figure C-1. Estimated NH$_3$-N Concentrations with Increasing VSR.](image-url)
Figure C-2. Treated Final Effluent NH₃ as a Function of Recycle NH₃ as a Percent of the Total Plant Load.
# APPENDIX D

## CONSTANTS FOR VSR MODEL CALIBRATION

Table D-1. Determination of the Hypothetical Sludge Retention Time SRT\(_1\) for a Maximum of Two Digesters.

<table>
<thead>
<tr>
<th>Process condition</th>
<th>(\beta) (days)</th>
<th>SRT(_1 \leq \beta)</th>
<th>SRT(_1 &gt; \beta)</th>
<th>SRT(_1 \leq \beta)</th>
<th>SRT(_1 &gt; \beta)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mesophilic digester</td>
<td>(\beta_M)</td>
<td>SRT(_1)</td>
<td>(\beta_M)</td>
<td>(\beta_M - SRT_1)</td>
<td>Negligible</td>
</tr>
<tr>
<td>Thermophilic digester</td>
<td>(\beta_T)</td>
<td>SRT(_1)</td>
<td>(\beta_T)</td>
<td>(\beta_T - SRT_1)</td>
<td>Negligible</td>
</tr>
<tr>
<td>Acid phase only</td>
<td>(\beta_A)</td>
<td>SRT(_1)</td>
<td>(\beta_A)</td>
<td>(\beta_A - SRT_1)</td>
<td>Negligible</td>
</tr>
</tbody>
</table>

SRT\(_1\) = Design SRT of stage 1; SRT\(_2\) = Design SRT of stage 2; SRT\(_1^*\) = Hypothetical SRT applicable to stage 1; SRT\(_2^*\) = Hypothetical SRT applicable to stage 2.

Table D-2. First Order Hydrolysis Constants and First Order Substrate Degradation Constants for PS and WAS.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Primary sludge</th>
<th>Waste activated sludge</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>First order hydrolysis constant ((K_{H,T,c,i}))</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mesophilic digester</td>
<td>0.150</td>
<td>0.120</td>
</tr>
<tr>
<td>Thermophilic digester</td>
<td>0.265</td>
<td>0.205</td>
</tr>
<tr>
<td>Pure acid phase digester (a)</td>
<td>0.460</td>
<td>0.190</td>
</tr>
<tr>
<td><strong>First order substrate degradation constant ((K_{1,T,c,i}))</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mesophilic digester</td>
<td>0.182</td>
<td>0.182</td>
</tr>
<tr>
<td>Thermophilic digester</td>
<td>0.390</td>
<td>0.390</td>
</tr>
<tr>
<td>Pure acid phase digester (a)</td>
<td>0.390</td>
<td>0.390</td>
</tr>
</tbody>
</table>

\(a\)Value derived for pure acid phase operating at mesophilic temperature.

Table D-3 Values of \(\beta\) for the Determination of Hypothetical Sludge Retention Times According to the Basic Processes.

<table>
<thead>
<tr>
<th>Process condition</th>
<th>(\beta) (days)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mesophilic digester (\beta_M)</td>
<td>21</td>
</tr>
<tr>
<td>Thermophilic digester (\beta_T)</td>
<td>10</td>
</tr>
<tr>
<td>Pure acid phase digester (a) (\beta_A)</td>
<td>6</td>
</tr>
</tbody>
</table>

\(a\)Value derived for pure acid phase operating at mesophilic temperature.
## PREDICTED SLUDGE TEMPERATURES IN U.S. AND CANADA

Table E-1. Average Sludge Temperatures in US States using Equation 5-18 (Chapter 5.0).

<table>
<thead>
<tr>
<th>State</th>
<th>Average annual temperature</th>
<th>State</th>
<th>Average annual temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>ambient°F °C</td>
<td>sludge°F °C</td>
<td></td>
</tr>
<tr>
<td>Alabama</td>
<td>62.8 17.1</td>
<td>69.1 20.6</td>
<td>Missouri</td>
</tr>
<tr>
<td>Alaska</td>
<td>31.5 -0.3</td>
<td>50.0 10.0</td>
<td>Montana</td>
</tr>
<tr>
<td>Arizona</td>
<td>60.4 15.8</td>
<td>67.6 19.8</td>
<td>Nebraska</td>
</tr>
<tr>
<td>Arkansas</td>
<td>60.4 15.8</td>
<td>67.6 19.8</td>
<td>Nevada</td>
</tr>
<tr>
<td>California</td>
<td>59.5 15.3</td>
<td>67.1 19.5</td>
<td>New Hampshire</td>
</tr>
<tr>
<td>Colorado</td>
<td>45.3 7.4</td>
<td>58.5 14.7</td>
<td>New Jersey</td>
</tr>
<tr>
<td>Connecticut</td>
<td>49.1 9.5</td>
<td>60.8 16.0</td>
<td>New Mexico</td>
</tr>
<tr>
<td>Delaware</td>
<td>55.4 13.0</td>
<td>64.6 18.1</td>
<td>New York</td>
</tr>
<tr>
<td>District of</td>
<td>54.4 12.4</td>
<td>63.9 17.7</td>
<td>North Carolina</td>
</tr>
<tr>
<td>Columbus</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Florida</td>
<td>70.7 21.5</td>
<td>73.8 23.2</td>
<td>North Dakota</td>
</tr>
<tr>
<td>Georgia</td>
<td>63.5 17.5</td>
<td>69.4 20.8</td>
<td>Ohio</td>
</tr>
<tr>
<td>Hawaii</td>
<td>75.6 24.2</td>
<td>76.8 24.9</td>
<td>Oklahoma</td>
</tr>
<tr>
<td>Idaho</td>
<td>44.4 6.9</td>
<td>57.9 14.4</td>
<td>Oregon</td>
</tr>
<tr>
<td>Illinois</td>
<td>51.8 11.0</td>
<td>62.4 16.9</td>
<td>Pennsylvania</td>
</tr>
<tr>
<td>Indiana</td>
<td>51.8 11.0</td>
<td>62.4 16.9</td>
<td>Rhode Island</td>
</tr>
<tr>
<td>Iowa</td>
<td>47.8 8.8</td>
<td>59.9 15.5</td>
<td>South Carolina</td>
</tr>
<tr>
<td>Kansas</td>
<td>54.3 12.4</td>
<td>63.9 17.7</td>
<td>South Dakota</td>
</tr>
<tr>
<td>Kentucky</td>
<td>55.6 13.1</td>
<td>64.6 18.1</td>
<td>Tennessee</td>
</tr>
<tr>
<td>Louisiana</td>
<td>66.4 19.1</td>
<td>71.2 21.8</td>
<td>Texas</td>
</tr>
<tr>
<td>Maine</td>
<td>41.0 5.0</td>
<td>55.8 13.2</td>
<td>Utah</td>
</tr>
<tr>
<td>Maryland</td>
<td>54.3 12.4</td>
<td>63.9 17.7</td>
<td>Vermont</td>
</tr>
<tr>
<td>Massachusetts</td>
<td>47.8 8.8</td>
<td>59.9 15.5</td>
<td>Virginia</td>
</tr>
<tr>
<td>Michigan</td>
<td>44.4 6.9</td>
<td>57.9 14.4</td>
<td>Washington</td>
</tr>
<tr>
<td>Minnesota</td>
<td>41.2 5.1</td>
<td>55.9 13.3</td>
<td>West Virginia</td>
</tr>
<tr>
<td>Mississippi</td>
<td>63.3 17.4</td>
<td>69.4 20.8</td>
<td>Wisconsin</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Wyoming</td>
</tr>
</tbody>
</table>
Table E-2. Average Sludge Temperature in Canada using Equation 5-18 (Chapter 5.0).

<table>
<thead>
<tr>
<th>Province</th>
<th>Annual average ambient temperature°</th>
<th>Average sludge temperature°</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>°F</td>
<td>°C</td>
</tr>
<tr>
<td>Alberta</td>
<td>37.0</td>
<td>2.8</td>
</tr>
<tr>
<td>British Columbia</td>
<td>44.4</td>
<td>6.9</td>
</tr>
<tr>
<td>Manitoba</td>
<td>35.1</td>
<td>1.7</td>
</tr>
<tr>
<td>New Brunswick</td>
<td>40.3</td>
<td>6.9</td>
</tr>
<tr>
<td>Newfoundland</td>
<td>38.5</td>
<td>3.6</td>
</tr>
<tr>
<td>Nova Scotia</td>
<td>43.5</td>
<td>6.4</td>
</tr>
<tr>
<td>Ontario</td>
<td>42.3</td>
<td>5.7</td>
</tr>
<tr>
<td>Prince Edward Island</td>
<td>41.9</td>
<td>5.5</td>
</tr>
<tr>
<td>Quebec</td>
<td>37.8</td>
<td>3.2</td>
</tr>
<tr>
<td>Saskatchewan</td>
<td>36.3</td>
<td>2.4</td>
</tr>
</tbody>
</table>
### APPENDIX F

## CONVERSION FACTORS

**Table F-1. Unit Conversions.**

<table>
<thead>
<tr>
<th>Convert From</th>
<th>To</th>
<th>Multiply by</th>
<th>Cautions and Assumptions</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Energy</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>British thermal unit (BTU)</td>
<td>Mega joule (MJ)</td>
<td>0.001054</td>
<td>Assuming thermal BTU</td>
</tr>
<tr>
<td>Kilo joule (kJ)</td>
<td>MBTU</td>
<td>0.000948452</td>
<td>Assuming thermal BTU</td>
</tr>
<tr>
<td>Kilowatt hour (kWh)</td>
<td>Mega joule (MJ)</td>
<td>3.6</td>
<td></td>
</tr>
<tr>
<td>Mega joule (MJ)</td>
<td>Kilowatt hour (kWh)</td>
<td>0.277778</td>
<td></td>
</tr>
<tr>
<td>Mega joule (MJ)</td>
<td>British thermal unit (BTU)</td>
<td>948.451653</td>
<td></td>
</tr>
<tr>
<td>MBTU</td>
<td>Kilo joule (kJ)</td>
<td>1054.35</td>
<td></td>
</tr>
<tr>
<td>MMBTU</td>
<td>British thermal unit (BTU)</td>
<td>1000</td>
<td></td>
</tr>
<tr>
<td><strong>Length</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Foot (ft)</td>
<td>Meter (m)</td>
<td>0.3048</td>
<td></td>
</tr>
<tr>
<td>Inch</td>
<td>Millimeter (mm)</td>
<td>25.4</td>
<td></td>
</tr>
<tr>
<td>Meter (m)</td>
<td>Foot (ft)</td>
<td>3.28084</td>
<td></td>
</tr>
<tr>
<td>Millimeter (mm)</td>
<td>Inch</td>
<td>0.0393708</td>
<td></td>
</tr>
<tr>
<td><strong>Mass</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Kilogram (kg)</td>
<td>Pound (lb)</td>
<td>2.204622</td>
<td></td>
</tr>
<tr>
<td>Kilogram (kg)</td>
<td>Short (ton)</td>
<td>0.001102311</td>
<td>Assume US (or short) ton</td>
</tr>
<tr>
<td>Kilogram (kg)</td>
<td>Metric tonne (MT)</td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>Mega gram (Mg)</td>
<td>Metric tonne (MT)</td>
<td>1</td>
<td></td>
</tr>
<tr>
<td>Metric tonne (MT)</td>
<td>Kilogram (kg)</td>
<td>1,000</td>
<td></td>
</tr>
<tr>
<td>Pound (lb)</td>
<td>Kilogram (kg)</td>
<td>0.4535924</td>
<td></td>
</tr>
<tr>
<td>Short ton (ton)</td>
<td>Kilogram (kg)</td>
<td>907.1847</td>
<td></td>
</tr>
<tr>
<td><strong>Mass per volume</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>Pound per cubic foot (lb/ft³)</td>
<td>0.062428</td>
<td></td>
</tr>
<tr>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>mg/L</td>
<td>1,000</td>
<td></td>
</tr>
<tr>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>Percentage (%) solids</td>
<td>0.1</td>
<td></td>
</tr>
<tr>
<td>Milligram per liter (mg/L)</td>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>Milligram per liter (mg/L)</td>
<td>Percentage (%) solids</td>
<td>0.0001</td>
<td></td>
</tr>
<tr>
<td>Percentage (%) solids</td>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>10</td>
<td></td>
</tr>
<tr>
<td>Percentage (%) solids</td>
<td>Milligram per liter (mg/L)</td>
<td>10,000</td>
<td></td>
</tr>
<tr>
<td>Pound per cubic foot (lb/ft³)</td>
<td>Kilogram per cubic meter (kg/m³)</td>
<td>16.01846</td>
<td></td>
</tr>
</tbody>
</table>
Table F-1. Unit Conversions (cont'd)

<table>
<thead>
<tr>
<th>Convert From</th>
<th>To</th>
<th>Multiply by</th>
<th>Cautions and Assumptions</th>
</tr>
</thead>
<tbody>
<tr>
<td>Power</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Horse power (hp)</td>
<td>Kilo watt (kW)</td>
<td>0.746</td>
<td></td>
</tr>
<tr>
<td>Joule per second (J/s)</td>
<td>Kilo watt (kW)</td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>Kilo watt (kW)</td>
<td>Horse power (hp)</td>
<td>1.34048</td>
<td>Assume motor/electric/ international hp</td>
</tr>
<tr>
<td>Kilo watt (kW)</td>
<td>Joule per second (J/s)</td>
<td>1000</td>
<td></td>
</tr>
<tr>
<td>Kilo watt (kW)</td>
<td>Mega joule per day (MJ/day)</td>
<td>86.4</td>
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</tr>
<tr>
<td>Mega joule per day (MJ/day)</td>
<td>Kilo watt (kW)</td>
<td>0.011574</td>
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</tr>
<tr>
<td>bhp(brake hp) = net available power(already includes engine efficiency)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>ie. bhp (brake hp) = hp x efficiency</td>
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<td></td>
<td></td>
</tr>
<tr>
<td>Power - Boilers Only</td>
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<td></td>
<td></td>
</tr>
<tr>
<td>Boiler horse power (bhp)</td>
<td>British thermal unit per hour (BTU/hr)</td>
<td>33472.119</td>
<td></td>
</tr>
<tr>
<td>Boiler horse power (bhp)</td>
<td>hp (electric)</td>
<td>13.15</td>
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</tr>
<tr>
<td>Boiler horse power (bhp)</td>
<td>Kilo watt (kW)</td>
<td>9.8097</td>
<td>not same as brake hp</td>
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<tr>
<td>British thermal unit per hour (BTU/hr)</td>
<td>Kilo watt (kW)</td>
<td>0.00029307</td>
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</tr>
<tr>
<td>British thermal unit per hour (BTU/hr)</td>
<td>Mega joule per day (MJ/day)</td>
<td>0.02532</td>
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</tr>
<tr>
<td>Kilo watt (kW)</td>
<td>British thermal unit per hour (BTU/hr)</td>
<td>3412.14245</td>
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</tr>
<tr>
<td>Temperature</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Degree Celsius (°C)</td>
<td>Degree Fahrenheit (°F)</td>
<td>1.8(°C) + 32</td>
<td></td>
</tr>
<tr>
<td>Degree Fahrenheit (°F)</td>
<td>Degree Celsius (°C)</td>
<td>(°F -32)/1.8</td>
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</tr>
<tr>
<td>Volume</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Cubic foot (ft³)</td>
<td>Cubic meter (m³)</td>
<td>0.02831685</td>
<td></td>
</tr>
<tr>
<td>Cubic foot (ft³) Natural Gas</td>
<td>British thermal unit (BTU)</td>
<td>1.000</td>
<td></td>
</tr>
<tr>
<td>Cubic meter (m³)</td>
<td>Cubic foot (ft³)</td>
<td>35.31466</td>
<td>Assume US liquid gallon</td>
</tr>
<tr>
<td>Cubic meter (m³)</td>
<td>Gallon (gal)</td>
<td>264.172</td>
<td></td>
</tr>
<tr>
<td>Cubic meter (m³)</td>
<td>Liter (L)</td>
<td>1.000</td>
<td></td>
</tr>
<tr>
<td>Gallon (gal)</td>
<td>Cubic meter (m³)</td>
<td>0.003785</td>
<td></td>
</tr>
<tr>
<td>Liter (L)</td>
<td>Cubic meter (m³)</td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>Mcf = Mscf = kscf</td>
<td>Cubic foot (ft³)</td>
<td>1.000</td>
<td>scf = standard ft³</td>
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<tr>
<td>MMcf = MMscf</td>
<td>Cubic foot (ft³)</td>
<td>1,000,000</td>
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</tr>
<tr>
<td>10 therms</td>
<td>British thermal unit (BTU)</td>
<td>1,000,000</td>
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</tr>
<tr>
<td>1 therm</td>
<td>Cubic foot (ft³) Natural Gas</td>
<td>100</td>
<td>approximate</td>
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### Table F-1. Unit Conversions (cont’d)

<table>
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<th>Multiply by</th>
<th>Cautions and Assumptions</th>
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</thead>
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<tr>
<td>Cubic foot per day (ft³/d)</td>
<td>Cubic meter per day (m³/d)</td>
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<td></td>
</tr>
<tr>
<td>Cubic meter per day (m³/d)</td>
<td>Cubic foot per day (ft³/d)</td>
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<td></td>
</tr>
<tr>
<td>Cubic meter per day (m³/d)</td>
<td>Gallon per day (gpd)</td>
<td>264.172</td>
<td>Assume US liquid gallon</td>
</tr>
<tr>
<td>Cubic meter per day (m³/d)</td>
<td>Mega gallon per day (Mgd)</td>
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</tr>
<tr>
<td>Gallon per day (gpd)</td>
<td>Cubic meter per day (m³/d)</td>
<td>0.003785</td>
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</tr>
<tr>
<td>Mega gallon per day (MGD) or 1,000,000 gpd</td>
<td>Cubic meter per day (m³/d)</td>
<td>3785</td>
<td></td>
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</table>

Digital Dutch, 2005
Estes, 1981
NW Natural, 2005
Online Conversion.com, 2005
Water Environment Federation, 2005
WebStat, 2005
REFERENCES


Liang, KY; Li, R; Tudman, S; Schneider, RJ; Sheehan, JF; and Anderson, E. 1999. Pilot Testing Case Study: Removal of volatile methylsiloxanes from Anaerobic Digester Gas Fired Engines. Presented at Annual Air and Waste Management (AWMA) Conference, St. Louis, Missouri.


Pohland, F.G. 1962. General Review of the Literature on Anaerobic Sewage Sludge Digestion. Engineering Extension Series No. 110, Purdue University, Lafayette, IN.


Cost-effective Energy Recovery from Anaerobically Digested Wastewater Solids


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Life Cycle Assessment Manager for Energy Recovery (LCAMER)

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LCAMER is intended to provide a consistent framework for analyzing the various factors and options available to wastewater treatment plants for recovering energy from the biogas produced during anaerobic digestion of wastewater solids. It provides a consistent and technically sound methodology to identify and analyze various design and operating conditions, and financial variables, which can affect decisions to adopt one or more digestion and energy recovery options or scenarios. Based on the inputs used in the tool, results are presented in terms of payback periods, so that different processes may be objectively compared, and decisions made regarding economic viability. Although the individuals (“Project Team”) who completed the work on behalf of the Sponsor made an extensive effort to assess the quality of LCAMER for consistency and accuracy, the LCAMER tool and/or any analysis results are provided on an “AS-IS” basis and use of LCAMER, the data information, or use of any method, or process as provided in LCAMER is at the user’s sole risk.

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Cobb County Water System
Colombus Water Works
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Maryland
Anne Arundel County Bureau of Utility Operations
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Wyoming, City of

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Missouri
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Orange Water & Sewer Authority

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Butler County Department of Environmental Services
Columbus, City of
Metropolitan Sewerage District of Greater Cincinnati
Northeast Ohio Regional Sewer District
Summit, County of

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Oklahoma City Water & Wastewater Utility Department
Tulsa, City of

Oregon
Clean Water Services
Eugene, City of
Gresham, City of
Portland, City of
Water Environment Services

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University Area Joint Authority

South Carolina
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Mount Pleasant Waterworks & Sewer Commission
Sparranburg Water

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Knoxville Utilities Board
Murfreesboro Water & Sewer Department
Nashville Metro Water Services

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Dallas Water Utilities
Denton, City of
El Paso Water Utilities
Fort Worth, City of
Houston, City of
San Antonio Water System
Trinity River Authority

Utah
Salt Lake City Corporation

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Fairfax County
Hampton Roads Sanitation District
Henrico, County of
Hopewell Regional Wastewater Treatment Facility
Loudoun County Sanitation Authority
Lynchburg Regional WWTP
Prince William County Service Authority
Richmond, City of
Rivanna Water & Sewer Authority

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King County Department of Natural Resources
Seattle Public Utilities
Sunrise, Port of
Yakima, City of

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Kenosha Water Utility
Madison Metropolitan Sewerage District
Milwaukee Metropolitan Sewerage District
 Racine, City of
Shibayang Regional Wastewater Treatment
Wausau Water Works

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South Australian Water Corporation
Sydney Water Corporation
Water Corporation of Western Australia

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United Kingdom
Yorkshire Water Services Limited

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Monterey, City of
Sacramento, County of
San Francisco, City & County of
Santa Rosa, City of
Sunnyvale, City of
Colorado
Aurora, City of
Boulder, City of
Georgia
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Cedar Rapids Wastewater Facility
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GH
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Name
Title
Organization
Address
City State Zip Code Country
Phone Fax Email

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Method of Payment: (All orders must be prepaid.)

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Account No. Exp. Date

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Shipping & Handling:

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<th>Canada &amp; Mexico Add:</th>
<th>All Others Add:</th>
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<td>8.00</td>
<td>50% of amount</td>
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<td>$20.00</td>
<td>5.00</td>
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<td>50% of amount</td>
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<td>30.00</td>
<td>5.50</td>
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</tr>
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<td>Add 20% of order</td>
<td>Add 20% of order</td>
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</tbody>
</table>

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