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FINAL
REPORT

State of Science Report:
Energy and Resource Recovery from Sludge

Global Water
Research Coalition

STATE OF SCIENCE REPORT: ENERGY AND RESOURCE RECOVERY FROM SLUDGE

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Global Water Research Coalition: Global cooperation for the generation of water knowledge

GWRC is a non-profit organization that serves as the collaborative mechanism for water research. The product the GWRC offers its members is water research information and knowledge. The Coalition will focus in water supply and wastewater issues and renewable water resources: the urban water cycle.

The founding members of the GWRC are: the Awwa Research Foundation (US), CRC Water Quality and Treatment (Australia), EAWAG (Switzerland), Kiwa (Netherlands), Suez Environment – CIRSEE (France), Stowa – Foundation for Applied Water Research (Netherlands), PUB—Singapore, DVGW – TZW Water Technology Centre (Germany), UK Water Industry Research (UK), Veolia – Anjou Recherche (France), Water Environment Research Foundation (US), Water Research Commission (South Africa), Water Reuse Foundation and the Water Services Association of Australia.

These organizations are all in charge of a national research program addressing the different parts of the water cycle. They have provided the impetus, credibility, and initial funding for the GWRC. Each brings a unique set of skills and knowledge to the Coalition. Through its member organizations GWRC represents the interests and needs of 500 million consumers.

The Global Water Research Coalition is affiliated with the International Water Association (IWA). The GWRC was officially formed in April 2002 with the signing of the partnership agreement at the International Water Association 3rd World Water Congress in Melbourne. With the U.S. Environmental Protection Agency a partnership agreement was signed in July 2003.

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EXECUTIVE SUMMARY

There is general consensus among sanitary engineering professionals that municipal wastewater and wastewater sludge is not a "waste", but a potential source of valuable resources. The subject is a major interest to the members of the Global Water Research Coalition (GWRC). The GWRC is therefore preparing a strategic research plan related to energy and resource recovery from wastewater sludge. The initial focus of the strategy will be on sewage sludge as i.e. water reuse aspects have been part of earlier studies. The plan will define new research orientations for deeper investigation.

The current state of science (SoS) Report was prepared as the preliminary phase of GWRC's future strategic research plan on energy and resource recovery from sludge. The goal of this report is to provide to the GWRC members the current knowledge on energy and resource recovery from sludge. The report is not intended to define future research paths, but to provide the necessary information that will allow GWRC members to define research priorities and objectives of joint project activities at a follow-up workshop.

The term "resource" used in this report indicates all materials and/or products other than energy that can be recovered from sludge such as phosphorus, building material, etc. The report focused on:

- The international situation of energy and resource recovery from sludge,
- How the use of different sludge treatment processes affects the possibility of recovering energy and/or materials from the residual sludge,
- The influence of market and regulatory drivers on the fate of the sludge end-product,
- The feasibility of energy and resource recovery from sludge,
- The social, economic and environmental performance (triple bottom line or TBL assessment) of current alternatives technologies.

A review of the international situation (Chapter 3.0) of energy and resource recovery from sludge showed that Sweden and Japan are probably the most advanced countries in the area. Many other countries including, The Netherlands, United States (U.S.), United Kingdom (UK), Germany, New Zealand, China, Malaysia, etc. have also been implementing energy and resource recovery from sludge for many years.

The review of current knowledge, based on literature survey, revealed that many technologies are able to recover energy and/or resources from sludge (Chapters 4.0 and 5.0). The technologies can be divided into two main categories, established and emerging technologies. The established technologies are those existing at full-scale with commercial applications, as well as those that can potentially be commercialized. International cases studies of such technologies are provided in Chapter 6.0 of the report. The emerging technologies include those that have been demonstrated only at pilot-scale or bench (laboratory) scale.

Energy recovery technologies can be classified into sludge-to-biogas processes, sludge-to-syngas processes, sludge-to-oil processes and sludge-to-liquid processes. The technologies available for resource recovery include those to recover phosphorus, building materials, nitrogen, volatile acids, etc. Technical, capital cost, operating and maintenance (O&M) costs information available were documented to the extent possible for each technology. Possibilities of upgrading biosolids pellets produced from sludge as renewable source of inoculum for bio-hydrogen gas production and also recovering of bio-pesticides from sludge are new research areas that have also been identified during the literature survey.

Four market drivers (Chapter 4.0) were identified and discussed including (in no particular order of importance):

- Sustainability and environmental concerns, such as the threat of soil pollution, global warming and resource depletion;
- Rising energy costs and the need of more electricity and heat to operate the plants;
- Requirements for high quality of resources for industrial applications, such as calcium phosphate for the phosphate industry; and
- Regulation as factor stimulating the development of new technologies.

With the large number of technologies available, it can be concluded that it is technically feasible to recover energy and building materials from sludge. It also well established that a resource like phosphorus (P) can be recovered with efficiency of 60-70%, and possibly higher. Although P recovery on full-scale is a technically feasible option, operating practice is in the early stages because most of the technologies are still in development.

To be attractive, technologies for energy and resource recovery must be affordable and cost-effective. This is currently not always the case. Some projects have failed because of the high capital O&M costs of the technologies. Examples of such projects are certain phosphorus recovery and building material production processes.

The social acceptance of a technology depends on the types of inputs used and the outputs generated. The technologies involving use of chemicals are more likely to be rejected by the public. Chemical use may be required in certain processes, but they may not always be the best options in term of health protection and life cycle impacts (energy use and emissions during production and transportation). For example, most current technologies for P recovery are based on extraction with sulfuric acid, a highly corrosive and potentially harmful chemical. In addition, technologies with high potential for pollutant emissions, either upstream or on-site, are assumed to have less public acceptance. Technologies involving several process units are generally viewed as less desirable complex processes, which require material and energy for production, greater land consumption, and higher capital and O&M costs, than simpler processes.

A TBL assessment showed that in term of energy recovery overall sludge-to-biogas processes are the most suitable options. For phosphorus recovery, it appeared that the technologies using less harmful chemicals like lime are the best options. Thermal solidification for brick production appeared as a better option compared to slag and artificial lightweight aggregates production. This cursory TBL assessment could not evaluate all technologies in depth, and should be used as general guide rather than as a definitive review. Indeed, many key information requirements are missing for some of the technologies, leading to incomplete or subjective assessment. The limits of the assessment are discussed in the report. A summary section of the TBL assessment places the playing field adopted for this report (i.e., energy and resources recovery from wastewater solids) relative to the wider assessment of the global environmental considerations for the entire treatment of municipal wastewater.

Based on the science and technology reviewed a series of knowledge gaps were developed. These included:

- Energy balance,
- Capital and O&M costs,
- Quantity of raw material used and resources produced,
- Technologies for
  - P recovery from metal precipitates,
  - coagulant recovery and recycling,
- Recovery of products on an elemental basis (e.g. nitrogen, phosphorus, carbon, sulfur, water)
- Life cycle analysis of the technologies,
- Identification of carbon footprints and GHG emissions
- Social acceptance surveys
- Modeling energy and resource recovery technologies and
- Optimal pathways for sludge treatment.

Recommendations regarding the knowledge gaps were provided.

1.0 General Introduction

1.1 Introduction

The wastewater treatment industry has a mandate to protect the water environment. In fulfilling this mandate, wastewater treatment plants produce large quantities of residual solids, commonly referred to as sludge. In order to fulfill completely its mandate, the industry needs appropriate strategies and methods for efficient management of sludge.

Agricultural use of sewage sludge has been the traditional approach for municipal sludge disposal. This approach has been used for many years mainly for two reasons, the first being that sewage sludge contains useful nutrients for crops. Sewage sludge contains at least 25% of the phosphorus that is present in raw wastewater (Balmér, 2004). If chemical precipitation or enhanced biological P-removal is employed, up to 95% of the wastewater phosphorus can be concentrated in the sludge (Balmér, 2004). Phosphorus in raw wastewater originates from human excreta and consumption of detergents. The quantity from these two sources is estimated at about 2.5 g P per person per day (Stark, 2005). Estimations from Matthews (1983) indicate that sewage sludge could cover over 5% of phosphate fertilizer needs in the UK. The second reason contributing to the use of sewage sludge in agriculture is the low cost. This method is in many cases the most cost-effective alternative for sludge management (Balmér, 2004).

Sewage sludge is comprised of many different components. Although sludge contains useful nutrients, it also contains harmful constituents such as heavy metals and pathogenic microorganisms. Direct use of sewage sludge in agriculture thus is appropriate only if the concentrations of the harmful constituents and pathogens are below the limit required for agricultural application. Because this is not always the case, consequently, sludge application in agriculture is no longer well accepted in many countries by the public, farming organizations, and food industry.

Unlike most other industries, the incoming wastewater to a treatment facility is a source of renewable resources. A large part of those resources are transferred to the sludge produced during the treatment of the liquid phase. Those resources include organic carbon compounds, inorganic compounds and nutrients. Thus sludge management practices in municipal wastewater treatment plants (MWTPs) should focus on processes that avoid the risks due to use of harmful constituents, and that generate valuable products that can be re-used. For example, energy can be extracted from the organic carbon compounds while construction materials can be produced from the inorganic compounds. Nutrients such as phosphorus can also be extracted and used as fertilizer.

Shizas and Bagley (2004) showed experimentally that sewage contains ten times the energy needed to treat it. During wastewater treatment, most of the soluble organic compounds that contribute to this measured energy are mineralized during secondary treatment to carbon dioxide (by cell respiration), or synthesized into new cell matter (cell growth). A small less-degradable fraction of the soluble organics is discharged in the treated effluent. Thus, only part of the total energy in the raw sewage as measured by Shizas and Bagley (2004) is recovered from residual wastewater solids, as a methane-rich gas following anaerobic digestion, as a synthetic gas (syngas), or by thermal oxidation. New development in microbial fuel cell technologies (Liu et al., 2004; Love, 2007) are making progress in capturing the energy from liquid wastewater, but the research is in the early stage of development. While there is opportunity for the wastewater industry to meet it own energy demand in the future, only energy in the solids can be extracted with current technologies. This report evaluates the energy recovery options only from the residual wastewater solids, but the liquid treatment processes.

1.2 Focus and Purpose of Report

The price of energy is anticipated to rise into the foreseeable future. MWTPs are almost always net consumers of energy, primarily as electricity, with some natural gas and diesel fuel use. Energy is one of the largest components of operating and maintenance costs for MWTPs. Consequently, the cost of MWTPs operations is expected to rise with energy costs. As the regulations for treated effluent quality become more stringent, energy consumption by the treatment facilities must increase as well (Monteith et al., 2007).

Coupled with rising energy costs is the concern over the contribution of electricity generation to global warming (GW). The contributions of water and wastewater treatment to total national energy consumption are in the low unit percentages, e.g. 1-5%. Sustainable wastewater treatment, with a reduced carbon footprint (CF), is now becoming a goal of major interest. Such interest has shifted the view of municipal sewage from a waste to be treated and disposed of, to a resource that can be processed for recovery of energy, nutrients or other constituents.

Currently technologies exist offering possibilities to recover resources such as electricity and/or heat, phosphorous, building material, etc. from sludge. Feedstock in the form of lipid can even be extracted from sewage sludge for diesel oil production. Dufreche et al., (2007) estimated that if 50% of MWTPs in the US were outfitted for lipid extraction and trans-esterification, production of 0.7×10^6 m^3 of biodiesel per year would result.

The Global Water Research Coalition (GWRC), in conjunction with it members, is preparing a strategic research plan on energy and resource recovery from municipal sewage sludge. This plan is intended to define new research orientations for deeper investigations.

The first step of any research plan is a clear synthesis of what has been already achieved in the area of interest. The current state of science (SoS) report provided herein is the preliminary phase of the future strategic research plan of the GWRC on energy and resources recovery from sludge. The scope and content of the SoS report were defined during a meeting at the headquarters of the Water Environment Research Foundation (WERF) in Alexandria, VA on July 16 and 17, 2007.

The goal of this report is to provide to the GWRC and its members a summary of the current knowledge on energy and resources recovery from sewage sludge. In this report, both the established technologies and emerging technologies are reviewed. Examples of the most commonly-used strategies are described, supported with international cases studies. The term "resource" is used in the report to indicate all materials and/or products (e.g. phosphorous) that can be recovered from sludge other than energy.

1.3 Audience

The SoS report will be used as a background document for the GWRC Workshop on energy and resource recovery from municipal wastewater sludge, to be held February 20 and 21, 2008. Objectives of the Workshop for the GWRC members, who are the target audience for the report, include: 1) identifying the research needs and knowledge gaps in energy and resource recovery from sludge; 2) prioritization of research needs to address the knowledge gaps; and 3) the development of research concepts and proposals.

2.0 Playing Field and Boundaries

Considerable effort went into definition of the report boundaries at the GWRC Steering Committee meeting of July 16-17. In general there are two ways to recover energy and resources in wastewater treatment plants. Energy and resource can be recovered from the wastewater and from the residual solids (sludge) and process streams.

The report focuses on energy and resource recovery from the residual solids and associated process streams, but not from the water phase directly. Residual solids are produced by different types of wastewater processes. The associated process streams include sludge supernatants, return sludge flows and sidestreams from solids dewatering processes, such as filtrates or centrates, which may contain elevated concentrations of nutrients or other resource materials. The types of wastewater treatment facilities considered in the this work include biological treatment processes such as activated sludge and its many variations, including biological nutrient removal; fixed film processes such as trickling filters; integrated fixed film activated sludge processes involving synthetic support media, and lower technology processes, such as lagoons and wetlands. The report addresses only centralized treatment facilities, and not individual or household treatment units. The size of the treatment facilities include any that are tied into a collection system operated by utility management.

The recovery of energy from sludge generally involves a conversion to either biogas, synthetic gas (syngas) or oil. This raw energy is then converted to into a useful form, electricity, mechanical energy, and/or heat. Many technologies can convert the raw energy into electricity and/or heat, including boilers, microturbines, engine generators, steam and gas turbines, combined cycle turbines, Stirling engines, direct drive engines, molten carbonate and phosphoric acid fuel cells. Detailed operation, commercial sizes, efficiency, etc., for those technologies have been documented by Monteith et al. (2006). A comprehensive discussion of anaerobic digestion, digester gas pretreatment options, and heat and energy recovery technologies is provided in Appendix A.

Finally, it is recognized that the recovery of energy and resource from municipal sludge will be dictated by certain drivers, such as the market for the recovered products, or by regulations that limit the options for recovery of material or energy. Public perception and social awareness may also prove to be drivers that influence the choice of recovery alternatives. All those market drivers will be discussed in the report.

3.0 Current International Practices

3.1 Regulations

In sewage sludge management, regulations are applicable to both the sludge treatment and disposal, as well as the extraction of energy. Because most sewage sludges contain harmful substances, the final disposal is subjected to a number of regulations by authorities in each country at the state/province or federal level.

In the EU, sewage sludge is regulated the Directive 86/278/EEC. This directive seeks to encourage the use of sewage sludge in agriculture, and to regulate its use in order to prevent harmful effects on soil, vegetation, animals and humans. In practice, the EU sewage sludge directive prohibits the use of untreated sludge on agricultural land unless it is injected or incorporated into the soil. The directive is implemented in each member country of the union through national regulation. For example, in the UK, the directive is implemented through the sludge regulations established in 1989. All sewerage companies were required to produce a sludge disposal strategy as part of the business planning process for 2005-2010.

The EU is reportedly revising the sewage sludge directive in 2007 (Gov-UK, 2007). This revision will have little impact in countries where sludge use is much more stringent than the EU requirements. Currently, in some European countries, direct use of sewage sludge in agriculture is no longer an option. This includes countries like the Netherlands and Sweden. In the Netherlands, agricultural spreading of sewage sludge has been forbidden since 1995 (Roeleveld et al., 2004). In 1988, 35% of the sludge produced in Sweden was used in agriculture. But the Federation of Swedish Farmers recommended its members not to use sewage sludge after January 1, 1990 (Hultman et al. 1998; Hultman, 1999).

The new EU regulations may have a more significant impact in countries like Greece, Ireland, Italy, Luxembourg, Portugal and Spain because in those countries, the current national requirement on sludge use is similar to the EU requirement (Stark, 2004).

Authorities in Switzerland (not a member of the EU), have proposed a ban on the use of sludge in agricultural applications after 2005 (Hultman et al., 2003). As alternative treatment, the Swiss government recommended a complete shift to incineration of sludge (ED, 2002a).

In North America sewage sludge management is regulated by different jurisdictional levels. In Canada the use or disposal of biosolids (treated sewage sludge) is regulated by the provincial governments. As a result there are significant differences between the approaches used to manage biosolids from one Canadian province to another.

In the United States of America, biosolids regulations have been developed at the federal level by the United States Environmental Protection Agency (U.S. EPA). Enforcement of the regulations lies either with the U.S. EPA, or to those states to which enforcement authority has been delegated. In the regulations, two categories, designated Class A and Class B, determine whether biosolids can be applied to agriculture or not.

Class A biosolids contain a very low level of pathogens. To achieve Class A requirements, sewage sludge must undergo heating, digestion or increased pH to reduce pathogens to non- detectable levels. Biosolids derived from those treatment processes can be applied to agriculture without pathogen restrictions, if the constituent metal concentrations are not an issue. Class A biosolids can also be marketed to the public for application in gardens.

Class B biosolids have less stringent standards for treatment and contain low densities of pathogens that are within compliance limits. The requirements for Class B are to ensure that pathogens in biosolids have been reduced to levels that protect public health and the environment.

Energy recovery from sewage and sludge, while favored environmentally, may also be subject to certain regulations. For example, in Ontario, Canada, new electricity production from anaerobic digester gas depends on the total power generated at the location. If it is less than 25 MW, the project is classified as Category A, and does not need an Environmental Assessment to proceed. Larger projects are subject to a full Environmental Assessment. Anyone can request that the Minister of the Environment make a Category A project subject to the Environmental Assessment Act.

In the State of Minnesota U.S., wastewater treatment plants are required to obtain air emission permits from the Minnesota Pollution Control Agency depending on the Total Maximum Rated Capacity (TMRC) of their energy recovery system. For example, installation of engine-generators with a TMRC exceeding 6,000 kW, which were installed after June 19, 1984, require an air quality permit.

Selection of processes which recover energy from anaerobic digester gas may be affected by the combustion efficiency and emission of pollutants such as nitrogen oxides (NOx) In the U.S, under Clean Air Act Amendments, areas with excessive concentrations of ground level ozone and smog are designated as ozone non-attainment areas. In those areas low emission technologies such as micro turbines and fuel cells are most appropriate for biogas energy recovery (Monteith et al., 2006).

3.2 Status of Sludge Production

Annual sludge production data are reported inconsistently in the literature, making it difficult to compare national averages. Table 3-1 shows the amount of sewage sludge generated in different countries of the European Union (EU). Overall, the total quantity of sludge has increased in 8 years by 43%, to 7.78×10^6 from 5.44×10^6 metric tonnes (MT) of dry solids per year. This is equivalent to an average increase of 5.4% per year. The highest sludge producers are Germany and the UK, while the lowest producers are Ireland and Belgium. Some nations such as Austria, Finland and Denmark have the lowest increases in sludge production over the period 1992 – 2000. The overall increase of sludge production is the direct consequence of progressive implementation of the Urban Waste Water Treatment (UWWT) Directive 91/271/EEC in all nations of the union. The UWWT Directive established that by December

2000 wastewaters arising from all agglomerations of more than 15,000 population equivalent (P.E) should be treated with a secondary treatment process (Europa, 2007). With growth of the population and more stringent requirement for the treatment of sewage effluent, sewage sludge production is expected to further increase (Gov-UK, 2007).

Table 3-1. Sewage sludge production (1000 dry MT of solids) in European Union from 1992 to 2000 (Extracted from Dichtl, 2003).

Countries	Year			
	1992	**1995**	**1998**	**2000**
Belgium	59	78	113	131
Denmark	175	185	200	200
Germany	2,208	2,512	2,514	2,736
Spain	528	751	787	1,069
France	643	764	878	980
Ireland	37	40	43	100
Netherlands	324	366	381	401
Austria	190	190	196	196
Portugal	126	147	246	348
Finland	150	158	150	150
UK	998	1,158	1,193	1,470
Total	**5,438**	**6,349**	**6,701**	**7,781**

Detailed data for national sludge production are difficult to acquire. Considerable effort was expended in trying to capture and compare national sludge production at a more detailed level. This report provides data from three nations, namely the Netherlands, the United Kingdom and the United States. As can be determined from the entries in Tables 3-2 through 3-4, the availability of data from one country to another can be highly variable.

In the United States, based on a sewered population of 222,840,915, and a per capita sludge production rate of 29.2 kg dry solids per year, the estimated U.S. sludge production for comparison with the data in Table 3-1 is 6,507 1000 MT of dry solids per year. The total sludge production for the U.S. is thus of the same magnitude as the EU nations reported in Table 3-1.

Table 3-2. Current wastewater sludge management practices summary for the Netherlands.

General Information		Number	Reference
Population receiving centralized wastewater treatment		16,225,000	(1-1-2004) Stichting Rioned
Population equivalents [1]		22,674,742	Benchmark 2006
Number of public wastewater treatment facilities		367	Benchmark 2006
Number of public wastewater treatment facilities (distribution by flow treated in m^3/d)			
0.000 – 0.378 m^3/d	0.000 – 0.100 mgd	5	Benchmark 2006
0.379 - 3.785 m^3/d	0.101 – 1.000 mgd	123	Benchmark 2006
3.786 – 37.850 m^3/d	1.001 – 10.000 mgd	206	Benchmark 2006
37.851 – 378.500 m^3/d	10.001 – 100.000 mgd	32	Benchmark 2006
378.501 m^3/d and greater	100.001 mgd and greater	0	
Total Flow Treated (m^3/d)		5,121,313	Benchmark 2006
Level of treatment provided by numbers of plants			
Less than secondary (>30 mg/l TSS and BOD_5)		0	
Secondary Treatment ($\leq$30 mg/l TSS and BOD_5)		367	
Advanced Treatment (nutrient or other pollutant removal included)		290	Estimate based on 80% total
Sludge Data			
Annual mass sludge per capita (dry kilograms/person)[2]		13.39	Benchmark 2006
Mass sludge per volume treated wastewater (dry metric tons/m^3)		0.19	Benchmark 2006
Number of central sludge processing facilities[3]		122	Benchmark 2006
Number of facilities with anaerobic digesters		92	CBS 2005
Number of facilities with anaerobic digesters and gas utilization		92	CBS 2005
Methane produced by sludge digestion (million metric tons/year)		0.04	CBS 2005
Current sludge end uses (percent)			
Combined Disposal (Incineration and Landfills)		76	CBS 2005
Land application		0	CBS 2005
Reuse (not land application)		24	Co-burning granulates by Cement industry or energy plants
Energy Data			
Energy (US cents per kilowatt hour) –average		14.1	Benchmark 2006
Energy consumed by wastewater treatment nationwide (billion kWh/year)		0.665	Benchmark (0.66 * 10^9 kWh)
Percentage of digester methane produced used for energy recovery		Not available	
Percentage of national energy output used for wastewater treatment		0.07	CBS 2003
Electric energy produced by co-generation of sludge (MW/day)		Not available	
Heat produced by sludge digestion (MJ/day)		Not available	
Carbon footprint of domestic and industrial wastewater treatment facilities (Tg CO_2 equivalent / year)		Not available	

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[1] Number of taxpayers (1 household = 3 pe)

[2] after digestion and dewatering

[3] digestion and/or dewatering

Table 3-3. Current wastewater sludge management practices summary for the United Kingdom.

General Information	Number	Reference
Population receiving centralized wastewater treatment	59,476,150	
Population equivalent of industries receiving public wastewater treatment services (in population equivalents)	72,723,140	
Number of public wastewater treatment facilities	9,312	
Number of public wastewater treatment facilities distribution by flow treated in)		
0.000 – 0.378 m^3/d 0.000 – 0.100 mgd	NA	
0.379 - 3.785 m^3/d 0.101 – 1.000 mgd	NA	
3.786 – 37.850 m^3/d 1.001 – 10.000 mgd	NA	
37.851 – 378.500 m^3/d 10.001 – 100.000 mgd	NA	
378.501 m^3/d and greater 100.001 mgd and greater	NA	
Total Flow Treated (m^3/day)	15,905,753	Calc. from 5,805,600 megaL/yr
Level of treatment provided by numbers of plants		
Less than secondary (>30 mg/l TSS and BOD_5)	NA	
Secondary Treatment ($\leq$30 mg/l TSS and BOD_5)	NA	
Advanced Treatment (nutrient or other pollutant removal included)	NA	
Sludge Data		
Annual mass sludge per capita (dry kilograms/person)	20.18	
Mass sludge per volume treated wastewater (dry metric tons/m^3)	0.21	Calc from 1.2 million tons per annum
Number of central sludge processing facilities	NA	
Number of facilities with anaerobic digesters	NA	
Number of facilities with anaerobic digesters and gas utilization	NA	
Methane produced by sludge digestion (million metric tons/year)	NA	
Current sludge end uses (percent)		
Combined Disposal (Incineration and Landfills)	20	
Land application	62	
Reuse (not land application)	18	
Energy Data		
Energy (US cents per kilowatt hour) –average	11.63	Eurostat 14 July 2006 8.22€ for 100 KWh
Energy consumed by wastewater treatment nationwide (billion kWh/year)	3.7	3,680 GWh /annum
Percentage of digester methane produced used for energy recovery	NA	
Percentage of national energy output used for wastewater treatment	NA	
Electric energy produced by co-generation of sludge (MW/day)	NA	
Heat produced by sludge digestion (MJ/day)	NA	
Carbon footprint of domestic and industrial wastewater treatment facilities (Tg CO_2 equivalent / year)	4	4 million MT CO2/year

Table 3-4. Current wastewater sludge management practices summary for the United States.

General Information	Number	Reference
Population receiving centralized wastewater treatment (74.9 %)	222,840,915	USEPA 2004 CWNS database
Population equivalent of industries receiving public wastewater treatment services (in population equivalents)	NA	
Number of public wastewater treatment facilities	16,583	USEPA 2004 CWNS database
Number of public wastewater treatment facilities (distribution by flow treated in m^3/d)		
0.000 – 0.378 m^3/d 0.000 – 0.100 mgd	6,830	USEPA 2004 CWNS database
0.379 - 3.785 m^3/d 0.101 – 1.000 mgd	6,431	USEPA 2004 CWNS database
3.786 – 37.850 m^3/d 1.001 – 10.000 mgd	2,771	USEPA 2004 CWNS database
37.851 – 378.500 m^3/d 10.001 – 100.000 mgd	503	USEPA 2004 CWNS database
378.501 m^3/d and greater 100.001 mgd and greater	41	USEPA 2004 CWNS database
Total Flow Treated (m^3/d)	127,405,604	USEPA 2004 CWNS database (33,657 mgd)
Level of treatment provided by numbers of plants		
Less than secondary (>30 mg/l TSS and BOD_5)	40	USEPA 2004 CWNS database
Secondary Treatment ($\leq$30 mg/l TSS and BOD_5)	9,221	USEPA 2004 CWNS database
Advanced Treatment (nutrient or other pollutant removal included)	4,916	USEPA 2004 CWNS database
Sludge Data		
Annual mass sludge per capita (dry kilograms/person)	29.2	NEBRA July 2007 Rpt *A National Biosolids … Survey (6.51mT/yr)*
Mass sludge per volume treated wastewater (dry metric tons/m^3)	0.14	Calculated from 2004 CWNS data
Number of central sludge processing facilities	2,000	Estimate based on email from Bastian USEPA
Number of facilities with anaerobic digesters (>17.0 m^3/d treated flow only)	544	USEPA 2004 CWNS database
Number of facilities with anaerobic digesters (> 17.0 m^3/d) and gas utilization	106	USEPA 2004 CWNS database
Methane produced by sludge digestion (million metric tons/year)	0.799	*Inventory of US Greenhouse Gas Emissions and Sinks*, USEPA April 2007 p. 8-8
Current sludge end uses (percent)		
Combined Disposal (Incineration and Landfills)	45	NEBRA July 2007 Rpt
Land application	49	Ibid.
Reuse (not land application)	6	Ibid.
Energy Data		
Energy (US cents per kilowatt hour) – average (US range 4.76 – 19.26)	8.77	Energy Information Admin. July 11, 2007
Energy consumed by wastewater treatment nationwide (billion kWh/year)	21	*U.S. Wastewater Treatment FS* Center for Sustainable Systems 2004
Percentage of digester methane produced used for energy recovery	34	USEPA *CHP at POTWs* April 2007
Percentage of national energy output used for wastewater treatment	3	*Water & Wastes Digest* Mar 2007
Electric energy produced by co-generation of sludge (MW/day)	114	USEPA *CHP at POTWs* April 2007
Heat produced by sludge digestion (MJ/day)	9,868,000	USEPA *CHP at POTWs* April 2007
Carbon footprint of domestic and industrial wastewater treatment facilities (Tg CO_2 equivalent / year) (US data includes septic as well as centralized treatment)	33.4	*Inventory of US Greenhouse Gas Emissions and Sinks*, USEPA April 2007
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One bench-marking number of interest from these tables is the estimate of dry sludge solids produced per capita. The figures for the Netherlands, the United Kingdom and the United States are 13.4, 20.2, and 29.2 kg dry solids per capita, respectively. Although the range is relatively broad, the range of per capita sludge solids production is likely a complex mix of social practices with respect to sewering of wastes, engineering design practices, and levels of wastewater treatment.

Data obtained from another source (Jacobsen, 2006) characterize the per capita sludge production numbers for EU member nations for the period 1992-2005. The results are provided in Figure 3-1. Compliance with EU Urban Wastewater Directive 91/271/EC in the 1990s for implementation of secondary treatment levels caused increases in a number of the EU member nations. The figure suggests that most member countries have sludge production numbers in the range of 20-35 kg dry solids per capita.

While there is reasonable agreement between the sludge production figures for the UK in Table3-3 and for year 2005 in Figure 3-1 (20 and 25 kg/capita/year, respectively), there is greater deviation for the figures for the Netherlands (14 and 26 kg/capita/year for Table 3-3 and for year 2005 in Figure 3-1, respectively). Differences in the reported figures may be due to different census and sludge production estimates. Other causes are uncertain.

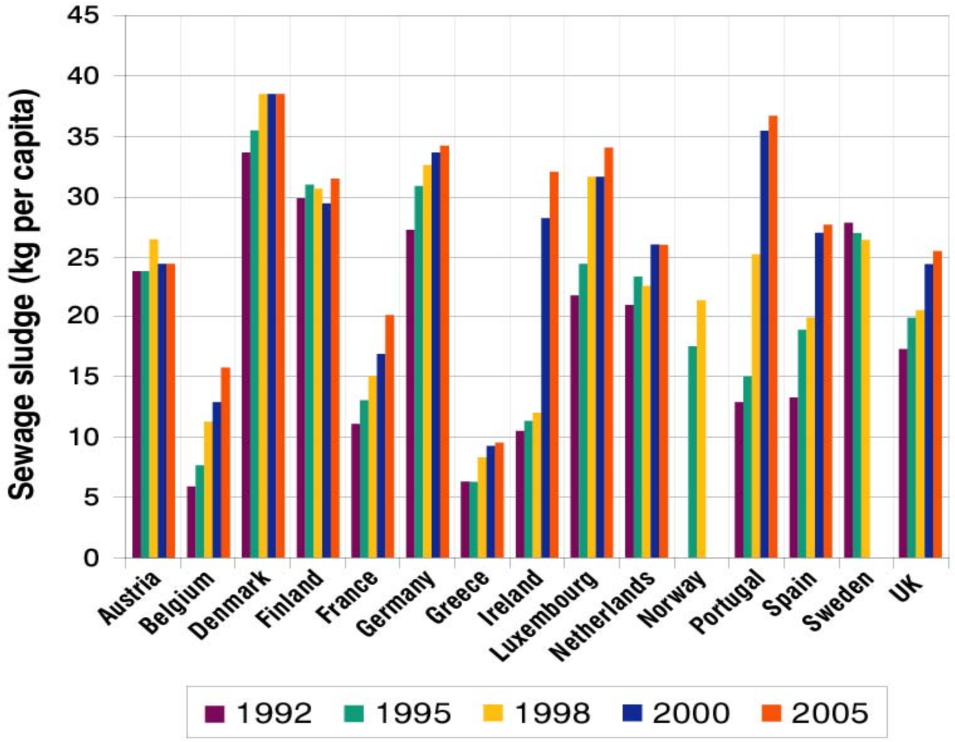

Figure 3-1. Per capita sewage sludge production in EU member nations, 1992 – 2005. (from Jacobsen, 2006)

Another benchmark of sludge production suggests less variability between the three countries with reported data in Table 3-2 through 3-.4. The mass of dry solids per volume of wastewater treated are reported as 0.19, 0.21 and 0.14 dry metric tones per m^3 of wastewater treated for the Netherlands, the United Kingdom and the United States, respectively.

Ultimate disposal of the sludge is another interesting comparison. Whereas the U.S. and U.K. rely heavily on land disposal (49% and 62%, respectively), the Netherlands has terminated the practice of land disposal due to potential health concerns. Instead, 76% of the sludge is incinerated, with the balance subject to reuse or resource recovery options.

3.3 Fate of Sludge End Product

3.3.1 Products Recoverable from Sewage Sludge

A large number of resources can be recovered from sewage sludge. Those resources include: nutrients such as phosphorus, nitrogen and potassium; organic compounds like organic acids, inorganic material, etc. Example of products that can be recovered and their end use are summarised in Table 3-5.

Table 3-5. Products recoverable from sewage sludge and their final use.

Type of Product	Use of product
Methane	Electricity, Heat, fuel
Gases	Electricity, Heat
Oil, fat, grease	Biodiesel, methane
Phosphorus	Fertilizers
Nitrogen	Fertilizers
Metals	Coagulants
Inorganic material	Building material
Organic compounds	Organic acid production
Inoculum	Bio-hydrogen gas production
Crystal proteins, spores	Bio-pesticides production

Kroiss (2004) assessed the potential for utilizing the nutrients in sewage sludge. He concluded that phosphorus is the most valuable compound in sewage sludge from the sustainability point of view but also in regard to the economic value. Firstly, this is due to the fact that the availability of phosphorus for the production of low cost mineral fertilizer is limited. Secondly, because nitrogen is an unlimited resource in the atmosphere from where, it can be recycled to agriculture via photosynthesis.

Sewage sludge is source of diverse inorganic compounds as can be seen by the composition of incinerated sludge ashes (Table 3-6). The content of ash depends mainly on the chemical used for coagulation of suspended particles during wastewater treatment (Levlin, 1999). The inorganic content of sludge can be used in useful and beneficial ways for the production various building materials.

Table 3-6. Range of element concentration in sludge ashes from various sources.

Parameters	**Netherlands** (Roeleveld et al. 2004)	**Japan** (Ozaki et al. 1997)	**Japan** (Takaoka et al. 1997)	**Sweden** (Stark, 2005)
Dry mass (%)	100	N.A.	N.A.	99.8
P_2O_5 (% dry solids)	±15-17	14.0-22.1	15.80	18.5
Al_2O_3 (%dry solids)	N.A	11.1-15.6	30.36	41.1
CaO (%dry solids)	N.A	7.6-12.3	12.94	6.91
Fe_2O_3 (% dry solids)	N.A	5.4-24.6	17.12	11.9
K_2O (% dry solids)	N.A	0.5-1.9	2.64	N.A
MgO (% dry solids)	N.A	2.5-3.8	1.05	0.99
Na_2O (% dry solids)	N.A	0.3-0.7	2.16	N.A
SiO_2 (% dry solids)	N.A	25.7-41.6	11.21	N.A
Fe (mg/kg)	80,000	N.A	N.A	N.A
Zn (mg/kg)	1,800	N.A	N.A	N.A
Cu (mg/kg)	1,100	N.A	130	N.A
Cd (mg/kg)	N.A	N.A	20	N.A
Pb (mg/kg)	N.A	N.A	206	N.A
Zn (mg/kg)	N.A	N.A	3,130	N.A
Ni (mg/kg)	N.A	N.A	546	N.A
Cr (mg/kg)	N.A	N.A	479	N.A
As (mg/kg)	N.A	N.A	8.2	N.A
Mn (mg/kg)	N.A	N.A	2,910	N.A

N.A Indicates information not available

3.3.2 Sludge Disposal Situation in Different Countries

3.3.2.1 Overview

Because of varied social and environmental values held by citizens, different countries may use different sludge disposal methods that reflect these values. For the member EU nations, Jacobsen (2006) has summarized graphically the different approaches to sludge disposal. This chart is provided as Figure 3-2. While expressed on a per capita sludge production basis, the figure indicates that certain European countries such as Denmark, Luxembourg and Finland rely heavily on recycling of sludge solids, other nations rely on other methods, such as incineration (the Netherlands), or landfilling (Greece).

There is now a consensus among experts in sanitary engineering that wastewater sludge is a source of valuable resources. Interest in extracting products from sludge, while not recent, is rising because of increases in energy costs, the threat of a decline in phosphate rock production, and impacts of global warming, to cite a few factors. Resource recovery from sludge is currently a worldwide topic and has become a key aspect of almost all sludge management master plans.

Sweden and Japan are probably the most advanced countries in the area of resource recovery from sewage sludge, based on the abundant literature available on practical and rewarding experiences in these two countries. Many other countries including the Netherlands, the U.S., the UK, Germany, New Zealand, China, and Malaysia have also implemented resource recovery from sludge for many years. Practical experiences in each country cited above are summarized in what follows.

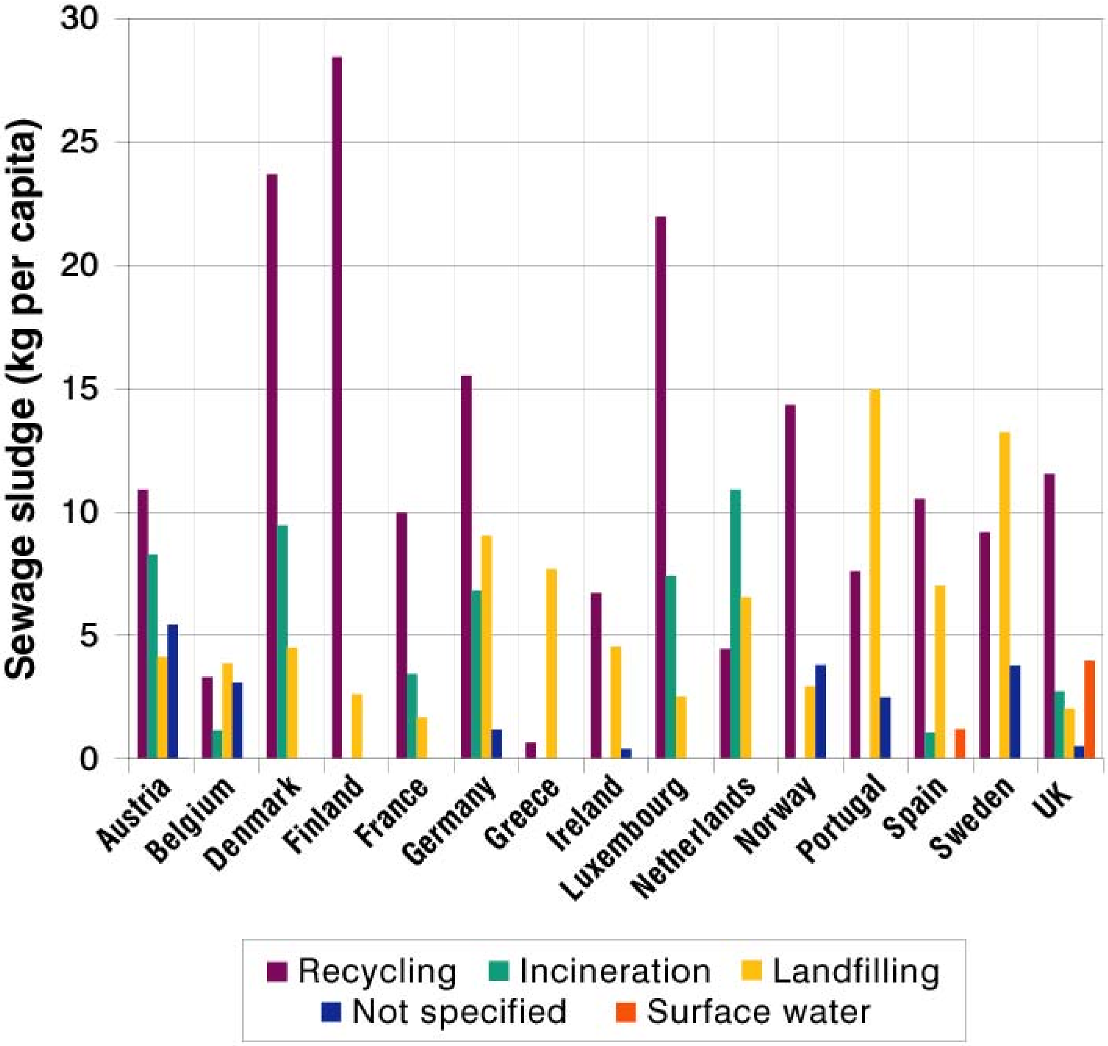

Figure 3-2. Sludge disposal practices in the European Union (Jacobsen, 2006).

3.3.2.2 Individual Countries

The United States of America

In the United States, beneficial use of biosolids on agricultural land (cropland, rangeland and pastures) is a major end-use. This may be in the form of liquid or dewatered anaerobically digested sludges, or as raw or digested sludges that have been treated by other methods, such as elevated pH by lime addition, or by aerobic composting, for reduction of pathogenic organisms and reduction of vector attraction. Heat-dried biosolids are also applied to land as soil conditioner, fertilizers, or fertilizers supplement (Sapienza et al., 2004). Additional nonagricultural uses include landfill cover, horticulture and silviculture, and reclamation of mine tailings.

OCEANGRO^{TM} is an example dried biosolids product used in the US for land application. This is a biosolid pellets produced by the Ocean County Utilities Authority. The product is utilized by over 60 New Jersey golf courses (Donovan and Ertle, 2007). In 2005 over three-quarters of the 9700 tons of OCEANGRO^{TM} produced was sold at an average price of approximately $25 per ton (Donovan and Ertle, 2007). The product is manufactures to fertilizer specifications and is registered with the New Jersey Department of Agriculture.

King County in Washington (Seattle) has recently investigated the use of biosolids produced by Seattle's two largest wastewater treatment plants as fertilizer to grow canola (Kris, 2007). This is a typical successful example of beneficial application of sewage sludge as resource for crops growth. Canola is a crop that can be used to make biodiesel fuel. In the King County operation, farmers crush the canola seeds and extract the oil. The farmers then sell the oil to a company called Imperium Renewables (Seattle), which processes it into biodiesel fuel. The city of Colton, California, conducted experiments to convert sewage sludge directly into biodiesel (VSA, 2006).

The city of Watsonville, California uses grease, delivered from restaurant grease trap haulers, to increase sewage sludge digester gas production by over 50%. The increase in biogas production offset the purchase of natural gas. The grease appears to digest completely and has little or no impact on downstream sludge dewatering and disposal processes (Cockrell, 2007). Grease from restaurants is a substrate suitable for anaerobic methane production because of the presence of energy rich compounds such as fats, carbohydrates, sugars, etc (Bailey, 2007).

Incineration is another major disposal technique. About 150 sewage sludge thermal oxidation installations were in operation in the US in 2002. Inert ash from the incinerators has been used in Columbus, OH, as a water-absorbent surface amendment in sport fields and horse arenas (U.S. EPA, 1994). Several municipalities, including Hampton Roads Sanitation District, Virginia and the City of Atlanta, Georgia have used have used the ash from thermal oxidation installations for brick production (Welp et al., 2002). Others, like MSD Cincinnati, Ohio have used ash for its nutrient value as a source of phosphorus (Welp et al., 2002). In the State of Minnesota, sewage sludge ash has long been used successfully in asphalt paving mixes and pavements (MWCC, 1990). According to (Donovan and Ertle, 2007) thermally dried biosolids is utilized in a cement kiln in Maryland. The biosolids are substitute for approximately 5 to 10% of the coal used to fuel the kiln.

There are several well established energy recovery technologies associated with sewage sludge treatment in the U.S. These have been documented in details by Monteith et al. (2006). Technologies include electricity and mechanical energy production, and heat recovery through methane generated from anaerobic digestion (McDannel and Wheless, 2007; Frankiewicz et al. 2007; Hake et al., 2007). The use of methane as source of hydrogen to produce energy with molten carbonate fuel has been demonstrated at King County, Washington's South Treatment Plant (Parry et al., 2004).

The U.S. Environmental Protection Agency (U.S. EPA) has instituted a program that seeks to reduce the environmental impact of electricity power generation. This program focuses on promoting the use of Combined Heat and Power (CHP) in different sectors including municipal wastewater treatment plants (MWTPs) (ERG and EEA, 2007). Those energy recovery systems include turbine, microturbine, fuel cell, reciprocating engine, etc. The program has been initiated because MWTPs that use anaerobic digesters to treat their waste sludges are an excellent technical fit for CHP (U.S. EPA, 2007). As an aid for municipalities considering energy recovery from digester gas, WERF funded a study from which the Life Cycle Assessment Manager for Energy Recovery (LCAMER) model was developed (Monteith et al., 2006). This model is a unique spreadsheet-based tool, which enables MWTPs owners and engineers to make informed decisions on the feasibility of recovering energy from anaerobic digestion of wastewater solids based on site-specific design and operating conditions, and energy pricing.

Under the EPA Environmental Technology Verification Program, different energy recovery technologies from biogas have been tested to evaluate their heat and power production performance, emission performance and power quality performance. Examples of technologies tested include a phosphoric acid fuel a cell and a microturbine (SRI 2004a, b).

The first commercial sludge-to-fuel facility is scheduled to open in mid-2008 in Rialto, California (Rozgus, 2007). This facility is expected to convert 675 wet tonnes per day of "E-fuel", which will be sold to a local cement kiln as a coal alternative. This project will make California the first state in the U.S. to power local industry with wastewater-treatment byproducts instead of coal.

The Netherlands

The Netherlands has been implementing phosphorus recovery from sewage sludge for many years. This country has stated the objective of replacing 20% of its current phosphate rock consumption by recovered phosphates (Roeleveld et al., 2004). Netherlands is recognized as one of the first countries that have implemented phosphorus recovery at full-scale in the Geestmerambacht MWTPs (Stark, 2004). In addition to recovered phosphorus, sludge is used as a make-up fuel. About 32% of the sewage sludge produced in Netherlands is currently used in cement industry and power stations (Uijterlinde, 2007).

United Kingdom

In the U.K., energy recovery from sewage sludge is a well-established practice. In 2005, waste (including sewage sludge) combustion and biogas production accounted for 10.8% and 4.2% respectively of all UK renewable energy (Trumper, 2007). Recently, a new program for energy recovery has been proposed by the government. This program includes strategies for generating 20% of electricity from renewable sources by 2020 (Trumper, 2007).

Experimental research was conducted to confirm the potential of sewage sludge as additive in brick manufacture (Anderson and Skerratt, 2003; Anderson 2002).

Germany

Germany is actively involved in the development of new technologies for P recovery from sewage sludge (Berg and Schaum, 2005). At least four technologies have been investigated either at pilot or bench scale since 2000. The technologies will be described later in this report.

Energy saving and recovery is also a priority in Germany. A renewable energy law was introduced by the German authorities in 1998 with the objective of funding the innovative and alternative technologies for energy production (Hannelore and Burkhardt, 2006). In 2002, the German Environmental Ministry funded the construction of a prototype sewage sludge incineration plant and power station in Sonthofen, Bavaria (ED, 2002b). This plant was built as a model that could be adopted elsewhere in the country if the trial results were successful.

Emschergenossenschaft (EG) and Lippeverband (LV) are publicly owned companies responsible for wastewater management in the catchments area of the rivers Emscher and Lippe. Currently, EG and LV are running intensive energy management programs for 59 MWTPs in order to optimize energy consumption (Kraft and Obenaus, 2007). The measures introduced at the wastewater treatment plants from 2003 to 2005 have reduced energy consumption by 3.2% and increased the onsite energy production by 27%.

Anaerobic co-fermentation of sewage and grease interceptor sludge is a method that is applied at the MWTP of Grevesmühlen to increase energy production (Schwarzenbeck et al., 2007). Grevesmühlen WWTP is located in the Northwestern part of Mecklengurg-West Pommerania and has a capacity of 40,000 population equivalents (P.E.). As a result of the co-fermentation, the plant produces excess electricity. In 2005 an average of 113% of the electricity consumed in the operation of the plant was generated on-site by gas engines. Co-fermentation of about 30% of grease interceptor sludge is reported to increase biogas production by a factor 4, and the anaerobic degradation of organic matter by 20% (Schwarzenbeck et al., 2007).

Sweden

Sweden is a country which has long understood that sewage sludge is a major source for biofuel. This country developed an impressive and strategic program of biofuel production from biogas. Practical examples of energy recovery from sewage sludge and effluent in Sweden have been reported by Salter (2006b) and Energie-cité (1999).

According to Salter (2006b), the Herenriksdales treatment plant produces and sells biogas to Stockholm's bus company. Biogas runs at least 30 buses in Stockholm. Stockholm's energy company (Fortum Energi) uses heat recovery pumps to extract heat from treated sewage effluent to provide hot water and heating to 80,000 apartments, including the Hammarby Sjöstad development. The sewage plant is paid for the thermal energy derived, which helps off-set the cost of treatment.

Biogas produced at other Swedish sewage plants is compressed and stored in a 25,000 kPa (250 bar) high-pressure vessel at a temperature of -30^{o}C (Energie-cité, 1999). It is then conveyed by truck to a centralized vehicle filling station. A special truck network was designed for the transportation of the biofuel. Figure3-3 shows an example of trucks used in Stockholm for biogas transportation to the gas filling station. In many other Swedish cities, including Trollhättan biogas produced from sewage sludge digestion is used as biofuel in the public transport network (Energie-cité, 2002).

In 2001, Sweden adopted an ambitious policy with a target of recovering and recycling 75% of phosphorus from sewage sludge by the year 2010 (Hultman et al., 2001; Levlin and Hutltman, 2003). Subsequent to the initial target, an intermediate target was established that by 2015, at least 60% of the phosphorus in wastewater was to be restored to productive soil (Hultman et al, 2003). Regulations have also been established requiring wastewater treatment plants to recover nutrients such as phosphorus when incineration of sludge is practiced (Hultman et al., 2001). As a consequence of this policy, new technologies for phosphorus recovery have been developed in Sweden. An extensive research program to improve the technologies is ongoing at the Kungl Tekniska Högskolan (KTH), the Swedish Royal Institute of Technology.

Figure 3-3. Example of truck used in Stockholm for biogas transportation to the gas filling station (Energie-cité, 1999).

China

In China, efforts have been made to use sewage sludge beneficially in bricks and other building material production (Wang, 1997). Methane fermentation is the most popular technology for energy recovery in China where the annual methane production from all feedstocks, including sewage sludge, was estimated at 720 million cubic meters (Aalbers, 1999).

New Zealand

The Council of the City of Dunedin in New Zealand, in July 2006 adopted a new waste management strategy based on resource recovery from solid, liquid and gaseous wastes (Dunedin, 2006). There are hopes that this plan will increase interest in resource recovery waste in the country.

The city of Napier is implementing the production of biosolids compost using primary sludge from the Awatoto Wastewater plant and wood chips (MacDonald et al., 2007). The goal of this project is to use the biosolids compost as soil conditioner and as a source of fuel for energy production.

Malaysia

In Malaysia research has been conducted to evaluate the potential of using dried sewage sludge as raw material to produce clay-sludge bricks (Liew et al., 2004a).

Japan

In Japan, many cities have considered the reuse of sewage sludge as a construction material and several municipalities and some companies have constructed plants in which construction material is produced form sewage sludge (Ozaki et al. 1997). Full-scale plants have been operating in Japan for more than a decade to produce construction material using inorganic material contained in sludge (Ozaki et al., 1997; Onaka, 2000; Okuno and Yamada, 2000; Okuno et al., 2004). Japan may therefore be considered as one of pioneer countries in production of construction materials from sewage sludge.

The Sewerage Bureau of Tokyo Metropolitan Government (SBTMG) has recently started a new project in which dewatered sewage sludge is turned into fuel charcoal and sold for thermal power generation (Oda, 2007). This is a new experience in Japan. Indeed, this is the first time this type of project is implemented in the country. The carbonization facilities which can treat 300 tons of dewatered sludge per day are currently under construction. The SBTMG has also developed a new system generating electricity with a gas engine using syngas produced by pyrolysis of sewage sludge (Takahashi, 2007). This system uses an internally circulating fluidized bed gasifier instead of the commonly used fluidized bed gasifier. The system is more efficient in term of emissions reduction and energy saving. As part of his effort for promoting energy recovery for sludge and reducing GHG emissions, the SBTMG has developed a non-conduit system to supply heat to multiple facilities such as hospital, hotels, sport centers, etc. up to a radius of 20 km (Yabuki and Nagumo, 2007). The project consists in distributing medium and low-temperature waste heat (300^{o}C to 70^{o}C) from sewage sludge incinerators. The heat is transported in containers with special trucks.

4.0 Review of Current Knowledge on Energy and Resource Recovery from Sludge

This chapter is the review of current knowledge (ROCK) on energy and resource recovery from sludge. The chapter is a synthesis of literature survey which examines:

- How the use of different sludge treatment processes affects the possibility of recovering energy or materials from the residual sludges,
- The influence of market and regulatory drivers on the fate of the sludge end-product,
- The feasibility of the applications and end uses of the sludge.

4.1 Categories of Treatment Processes for Resources Recovery

Physical, thermal, biological and chemical processes can be used for resource recovery from sludge. Table 4-1 summarizes the processes based on the type of recovered resource. Examples of technologies for each process are also presented in Table 4-1. In this chapter, only the established technologies are discussed. The emerging technologies for resource or energy recovery will be discussed in Chapter 5.0.

Table 4-1. Categories of treatment processes for resource recovery.

Processes	Example of Technology
Phosphorus	
Chemical processes	KREPO, Seaborne, Aqua-Reci, Kemicond, BioCon, SEPHOS
Crystallization processes	Crystalactor®, Phostrip
Building material	
Thermal solidification	GlassPack®
Incineration	Portland cement
Nitrogen	
Chemico-process	ARP Technology
Volatile acids	
Microbiological	Fermentation
Hydrothermal	Wet air oxidation

4.2 Phosphorus Recovery

4.2.1 Calcium Phosphate Recovery

There are two principal types of calcium phosphate recovery, namely those using a seed material for development of calcium phosphate crystals, and one involving calcium phosphate precipitation by lime addition to a phosphorus-rich sludge side-stream.

4.2.1.1 Seeded Process

Crystalactor® Technology

The Crystalactor® Technology uses sand as the seed material for crystal development. The process reactor is filled with sand as seeding material. In operation, the reactor receives a portion of the settled activated sludge from the secondary clarifier of an BNR plant. A solution of lime [$Ca(OH)_2$] is added to the reactor to increase both the pH, to about 8, and the concentration of calcium ions to create optimal conditions for precipitation of calcium phosphate. With time, the calcium pellets increase in size and weight. Larger and denser pellets descend to the bottom of the reactor. The pellets are discharged from the bottom of the reactor and make-up sand material is added. The pellets typically consist of 40 - 50% calcium phosphate, 30 - 40% sand, and up to 10% calcium carbonate (STOWA, 2006a). The Crystalactor® Technology has been applied at full-scale in the Netherlands. The technology was installed at the Geestmertambacht, Heemsted and Westerbrork (Stack, 2007). The cost of calcium phosphate production using the Crystalactor® process has estimated as 22 times higher than the cost of mined phosphate rock (Roeleved et al., 2004), and thus is not considered viable form a purely economical view.

The Crystalactor® process is an add-on, and does not require significant modification to existing sludge handling processes. The process requires readily available lime to raise pH. It does not need hydrocarbon fuel inputs, therefore has a lower carbon footprint. The product can be used as fertilizer raw material. Cost of production is high relative to natural sources, but the value of the recovered calcium phosphate should increase as natural supplies dwindle.

P-RoC

The P-RoC process uses surplus settled activated sludge in the same way as Crystalactor®. The seed material in the P-RoC or Phosphorus RecOvery process under development in Germany is a tobermorite-rich waste material from the construction industry (Berg and Shaum, 2005). No additional lime is required as the tobermorite is composed of Calcium silicate hydrates. Tobermorite appears to stimulate the precipitation of calcium phosphate, serving as the crystallization nucleus, while it also increases the reactor pH due to its chemical properties, reducing the solubility of the calcium phosphate. As the process is under development, no cost data are available.

P-RoC uses waste material as both the seed for crystal development and as the pH adjustor. It does not need large energy inputs, therefore has a lower carbon footprint. Product can be used as fertilizer raw material. Cost of production is unknown as no pilot- or full-scale implementation yet. The value of the recovered calcium phosphate should increase as natural supplies dwindle.

4.2.1.2 Precipitation

Phostrip© Technology

The Phostrip© Technology recovers calcium phosphate from a phosphorus-enriched sludge processing side-stream. While technically not applied to the sludge itself, the process is considered in this review be cause it treats the return activated sludge stream. A portion of the return sludge from the clarifier is pumped to an anaerobic stripper tank to release phosphorus to the liquid phase. Acetic acid may be dosed to the stripper to increase the amount of phosphorus released. After the phosphorus-rich water is separated from the sludge, it is treated with lime to precipitate the phosphorus as calcium phosphate.

PhoStrip© was installed in the 1970s in the Reno/Sparks Joint Water Pollution Control Plant located in Sparks, Nevada, U.S. (Levlin and Hultman, 2003). Since then, the technology has been used in many municipal wastewater treatment plants (MWTPs). At least 4 four full-scale plants were using the process in the mid-1980s, including Lansdale, PA; Little Patuxent, MD; Central Contra Costa Sanitation District, CA and Seneca Falls, NY. Design parameters of those plants were summarized by Rybicki (1997).

Phostrip process is an add-on, and does not require significant modification to existing BNR plants. The process requires readily available lime to raise pH and precipitate calcium phosphate. It does not need hydrocarbon fuel inputs, therefore has a lower carbon footprint. The product can be used as fertilizer raw material. Cost of production is high relative to natural sources, but the value of the recovered calcium phosphate should increase as natural supplies dwindle.

Struvite Formation

Struvite is crystalline magnesium ammonium phosphate hexahydrate $(NH_4)MgPO_4 \cdot 6(H_2O)$. The composition of struvite is approximately 30% by weight expressed as P_2O_5. Unintended formation of struvite in wastewater treatment is usually detrimental to operation (blocking pipes and fouling heat transfer surfaces).

Weidelener et al (2007) has proposed a method to leach phosphorus out of digested sewage sludge and produce struvite. This method uses sulfuric acid as a leaching agent. Prior struvite precipitation with $MgCl_2$, interfering metal ions in the leachate are inactivated through complexation. This method allows to produce a product comparable to mineral commercial fertilizers in term of heavy metals concentration.

The OSTARA process (Canada) recovers struvite from a phosphorus-rich sludge stream using magnesium chloride. Supplemental caustic soda may be required depending on the alkalinity and hardness of the P-bearing waste stream. A full-scale facility has been in operation at the City of Edmonton, AB's Clover Bar sludge processing facility. A portion (approximately 20-25%) of the supernatant from the sludge lagoon is the feed material for the struvite recovery. The process used a fluidized bed reactor to generate the struvite crystals. Phosphate recovery is on the order of 80-85%. The process has been operational since May 2007, and is expected to produce between 200-250 MT/year of struvite. Capital cost of the facility was approximately $2.5 million Cdn. Plans are underway to expand the facility to treat all of the supernatant, creating an annual yield of 1100 MT of struvite. Additional OSTARA process installations are in planning for Suffolk, VA and Portland, OR (Mavinic, D. 2007; Prasad et al., 2007).

A full-scale process has been installed to recover phosphorus as struvite at the Lake Shinji East Clean (LSEC) Center in Japan (Ueno and Fujii, 2001). The feed to the process is not wastewater sludge directly, but the phosphorus-enriched filtrate of a sludge dewatering process. The excess sludge from the plant's biological nutrient removal (BNR) process is digested anaerobically, releasing its phosphorus content. After dewatering of the digested sludge, the phosphorus-rich filtrate is enters the bottom of a precipitation reaction tower at a concentration of 100 to 110 mg/L as phosphoric acid. Magnesium hydroxide and sodium hydroxide are also injected to adjust the reactor pH between 8.2 and 8.8. Struvite crystals grow in the reactor to 0.5 to 1.0 mm after about 10 days. The produced struvite is recovered at the bottom of reactor and sold as fertilizer. The treated effluent leaves the top of the reactor with a phosphoric acid concentration of 10 mg/L, achieving more than 90% removal of phosphorus. With a treatment capacity of 45,000 m^3 per day, the LSEC produces approximately 500 to 550 kg of struvite per day, equivalent to approximately 0.01 kg struvite per m^3 of wastewater treated.

With the struvite formation processes, ammonia is present in the sludge, while chemicals magnesium and sodium hydroxide are easily obtained. Caustic soda is more hazardous to handle than chemicals used in some other P recovery processes using lime. Retrofit of process not too difficult, however anaerobic digestion of sludge must be in place. If the digested sludge is not dewatered, a mechanical dewatering device (electrical-energy intensive) is required Electricity is also required for the reactor and for separating the struvite from the residual material. Struvite market price is uncertain relative to cost to produce, but value is likely to rise as phosphorus supplies dwindle.

4.3 Building Material Recovery

Typical building materials that can be produced from sewage sludge include artificial lightweight aggregates (ALWA), brick, slag, ceramic, cement, glass, interlocking tile, etc. In general, the materials are produced after ash has been recovered following sludge incineration. For brick and cement production, however, the sludge can be used directly without incineration. The main process used to produce the material is thermal solidification. This process consists of melting and then solidifying the ash.

Processes used to produce ALWA, brick and slag have been described by Okuno et al. (2004). ALWA are produced by blending ash with water and a small amount of alcohol-distillation waste as a binder material. The mixture is supplied to a centrifugal pelletizer. The pellets produced are dried at 270°C for 7-10 minutes, with the dried pellets heated at 1050°C in a fluidized bed kiln for a few minutes to produce the final aggregate material.

Slag is a marble-like mineral of semi-crystalline structure. For production of slag material, sludge incinerator ash is initially blended with a small amount of lime. The slag is produced by melting ash in a cyclone furnace pre-heated to 1500°C. The slag can be used in the production of cement and mortar, or as an inert fill material.

The production of sludge-based bricks requires pre-treatment of the sludge ash at high pressure. This is followed by heat treatment at a temperature up to 1000°C. After the pressure and heat treatment, the material is finally poured into a die and pressed up to 1000 kg/cm^2.

Table 4-2 prepared from information reported by Okuno and Yamada (2000) and Onaka (2000) shows three plants and the processes used for producing building materials with sewage sludge in Japan.

Table 4-2. Example of plants producing building materials from sewage sludge in Japan.

Plant Name	Start-up	Unit Process	Product Name	Product Quantity	Products Use
Nambu [a]	1996	-Water/alcohol Blender -Centrifugal Pelletizer -Dryer(270°C) -Fluidized kiln(1,050°C)	ALWA	500 kg per h	-Thermal insulator -Water-infiltration plate -Fillers
Minami [b]	1990	-Steam dryer -Hot blast crusher -Melting furnace (1,500°C)	Slag	540 kg per h	-Concrete aggregates -Interlocking tiles -Water permeable tiles
Tagawa [b]	1998	-Pelletizing -Drying	Pellets	210 kg per h	-Portland cement

[a]Ash from sludge incineration is used as feed, [b]Dewatered sludge cake is used as feed

At Maastricht in the Netherlands, the industry ENCI uses sewage sludge as fuel in cement kilns (Ketton, 2004). The plant has completed a permit procedure for the use of certain categories of waste and biomass as secondary fuel (Würdemann and van Veen, 2002) .The ENCI cement-kilns employ a co-incineration process using sewage sludge along with agricultural waste and paper sludge as a fuel source.

Co-incineration of sewage sludge with municipal solid waste is applied in other countries like Sweden (Levlin, 1999). Sewage sludge can also be co-incinerated with coal. Co-incineration of sewage sludge with MSW or coal may have cost advantage over mono-incineration (sludge only) depending on supply and transport factors (BAT and BEP, 2004).

GlassPack® is a patented vitrification process, developed in North America by Minergy Corporation, which can be integrated into a wastewater treatment plant using biosolids as the feed material. The technology uses the organic fraction of biosolids as a renewable fuel source to produce an inert glass aggregate product from the inorganic (ash) fraction. Wet sludge, with 17 to 20% solids, is pre-dried to <15% moisture; the dried solids are then subjected to temperatures between 1,330 and 1,500°C, at which the ash component melts into molten glass. The molten glass and exhaust gas are separated by gravity draining of the glass into a quench tank. The gas is exhausted from the melting unit to a heat recovery system.

Since September 2006, the North Shore Sanitary District (NSSD) has used the GlassPack® vitrification process to treat 66,000 MT per year) of biosolids (18% dry solids) from their three wastewater treatment plants. The aggregate has multiple beneficial reuse outlets. The glass aggregate produced at NSSD is approved for beneficial reuse by the Illinois EPA and Wisconsin Department of Natural Resources. Local municipalities use the aggregate as fill in utility trenches (Minergy, 2007). The technology is further described in Chapter 6.0 as a Case Study.

Thermal solidification processes are energy intensive. Energy consumption for slag, brick and ALWA production are summarized in Table 4-3.

Table 4-3. Energy consumption for Slag, Brick and ALWA production (Adapted after Okuno & Yamada, 2000).

Product	Natural Gas (m^3/d MT cake)	Electricity (kWh/MT cake)	Total Energy [b] (kWh/MT cake)	Total Energy [c] (kWh/dry MT cake)
Slag	35.0	130	497.5	1,658
Brick-incineration ash [a]	40.4	206	630.2	2,101
ALWA [a]	38.1	156.5	556.7	1,856

[a] Include energy used during incineration step; [b] Sum of electricity and natural gas expressed in kWh; [c] assume 30% dry solids in dewatered sludge

4.4 Energy Recovery

The discussion in this report is focused on the recovery of energy from the solids handling processes, rather than the liquid treatment units. While the solids streams are enriched in organic solids, from which energy is extracted, it is noted in passing that the latent heat value of treated wastewater is a resource which is being tapped in certain countries for centralized heating, for example, in Sweden the cities of Stockholm and Gothenburg use heat pumps to recover thermal energy from the treated municipal wastewater (Salter, 2006).

Physical, mechanical, biological and chemical processes can be used to produce or contribute to energy recovery from sludge. These processes can be grouped into four main categories; Sludge-to-Biogas processes, Sludge-to-Syngas processes, Sludge-to-Oil processes and Sludge-to-Liquid. Examples of technologies (established and emerging) for each category are summarized in Table 4-4. This section discusses only the established technologies. The emerging technologies will be discussed in Chapter 5.0.

Table 4-4. Categories of treatment processes for energy recovery.

Processes	Example of Technology
Sludge-to-Biogas	
Anaerobic digestion	Bioterminator[24/85]
Thermal hydrolysis	Cambi®, BioThelys®,
Physical-chemical Cell destruction	MicroSludge™, Ultrasonic, Ozonation, Pulse electric
Sludge-to-Syngas	
Gasification	KOPF, EBARA
Incineration	Thermylis® HTFB
Sludge-to-Oil	
Pyrolysis	Enersludge™, SlurryCarb™
Hydrothermal	STORS
Sludge-to-Liquid[a]	
SCWO	Aqua Reci®, Aqua citrox®, Athos®

[a] Sludge is converted to a liquid form and heat is recovered

4.4.1 Sludge to Biogas

4.4.1.1 Sludge Pretreatment for Enhanced Biogas Production

A primary operational goal of wastewater treatment is to extract the maximum quantity of biogas possible from feed sludge. Several processes have been developed to break down the raw sludge solids to promote easier biotransformation to methane. Thermal, mechanical and chemical cell destruction processes have been developed, including hydrothermal heating, ultrasonic cell disintegration, use of ozone and electrical pulses.

Because these processes are associated with greater energy recovery as biogas from the anaerobic digestion process, they have been included as part of this review.

Thermal hydrolysis

Thermal hydrolysis is a process used to increase sludge digestibility for better biogas production, and to decrease the quantity of residue for disposal. During thermal treatment, sludge is heated at high temperature and high pressure for several minutes. Microbial cell walls in the sludge are destroyed, releasing more easily digestible organic compounds contained within the cells.

The advantages of combining thermal hydrolysis and anaerobic digestion are high volatile solids (VS) destruction and increased biogas production. Currently, the most known commercial thermal hydrolysis technologies are Cambi® and BioThelys®.

Cambi® Technology

Cambi® Technology consists of a series of three closed reactors, including a pulping vessel, a hydrolysis reactor and a flash tank (Steve and Panter, 2002). The Cambi® system involved batch treatment of both undigested primary and secondary sludge. Dewatered sludge (approximately 16 % solids) is pre-heated with steam to 80°C in the pulping vessel. The pre-heated sludge is sent to the reactor, operating at elevated temperature (160-180 °C) and pressure (600 kPa or 6 bar) for about 30 minutes. From the thermal reactor, the sludge is transferred to the flash tank, operating at atmospheric pressure. The change in pressure from the hydrolysis reactor to the flash tank causes cell lysis. The hydrolyzed sludge is then cooled for mesophilic anaerobic digestion.

Cambi® installations were identified in Norway, Denmark, England, Ireland, Scotland and Poland. At the HIAS MWTP in Hamar, Norway, biogas production rose by approximately 50% with operation of the Cambi® process over the production rate with anaerobic digestion alone (Kepp et al. 1999).

The technology is moderately complex with three reactors, one of which operates at elevated temperature and pressure. Solids from wastewater treatment must be dewatered prior to the process, and a medium-pressure steam supply is required. Reports of odor problems have been associated with the process.

A case study of Cambi® operation is presented in Section VI. A more detailed description of the process is found in Appendix B.

BioThelys® Process

The BioThelys® process, developed in France, uses a single reactor to treat a sludge feed with a solids concentration higher than 10% dry solids. The reactor operates at temperatures between 150°C and 180°C and pressures between 800 and 1000 kPa (8 and 10 bars). The retention time varies between 30 and 60 minutes. According to the vendor, this technology can reduce sludge production by up to 80%. Two full-scale facilities have been in operation in France since 1998. Additional details of operations and cost are not available.

Like the Cambi® process, the BioThelys® process requires high temperature and pressure for the cell hydrolysis, although only one reactor is required. While not identified, the process may also be subject to odor concerns.

A more detailed description of the process is found in Appendix B.

Physical-Chemical Cell Destruction

Cell destruction involves processes other than thermal hydrolysis, such as ultrasonic treatment, ozonation, pulse electric fields and mechanical disintegration to destroy the cellular membrane of the microorganisms in the sludge, resulting in release of the soluble cell contents. The ruptured sludge solids are then treated in an anaerobic digester. In anaerobic digestion alone, cell destruction requires more time than does hydrolysis of the soluble cell contents, and is thus the rate limiting step. With cell destruction, the membranes are lysed more quickly than in digestion alone, so the overall time for digestion is reduced. This results in more rapid and increased degradation of the volatile solids (VS) in the digester, reportedly up to 80-90%. Combination of a cell destruction process with anaerobic digestion produces more biogas than conventional anaerobic digestion alone. Examples of cell destruction technologies are MicroSludge™, ultrasonic treatment, ozonation, pulse electric fields and mechanical disintegration.

MicroSludge™

This technology mixes thickened waste activated sludge (5 to 10% dry solids) with caustic soda for about one hour to weaken cell membranes. High pressure (about 800 kPa or 80 bars) is next applied in a homogenizer, or cell disrupter, to provide an enormous and sudden pressure drop (from 800 to 40 kPa), which lyses the bacterial cells in the sludge. The liquefied sludge is then mixed with primary sludge and treated in an anaerobic digester with a reported detention time less than 9 days.

Microsludge is reported by the vendor to be a net energy producer, consuming only 37% of the biogas energy as electricity (30% electrical energy conversion). The process does not require dewatering of secondary solids, only thickening. Detention times in the downstream anaerobic digester can be shortened, while reportedly maintaining a high volatile solids reduction. Disadvantages include use of caustic soda, corrosive chemical for handling. High pressure is required for homogenization. Other potential problems include process odors, equipment complexities and high organic/nutrient loads back to the headworks (Toffey, 2004). The first full-scale MicroSludge^{TM} process was installed at the Chilliwack MWTP near Vancouver, BC, Canada in 2004, with a second full-scale demonstration installation started-up at the Los Angeles County Sanitation District (LACSD) in October, 2005. The technology is no longer in operation in those two plants. Based on the results of full-scale and bench-scale tests, the LACSD determined that the MicroSludge process was not cost effective for treating thickened waste activated sludge (TWAS) at the Joint Water Pollution Control Plant (JWPCP). The economic analysis of implementing this process at the JWPCP showed that the installed capital cost was estimated at $35 million per MGD of TWAS treated, and the anticipated O&M cost was approximately $2.5 million per year per MGD of TWAS.(Gary et al., 2007). The Chilliwack plant discontinued use of the process to re-evaluate the operation there following the JWPCP tests (Mavinic, 2007)

Ultrasonic Treatment

This technology applies ultrasonic acoustic waves to wastewater sludge to produce a very high pressure and temperature within the sludge. This results in cavitation, or the implosion of gas bubbles, which produces large shear stresses that break up membrane walls of bacteria and other cellular matter (Hogan et al., 2004). Ultrasonic treatment is typically applied to waste activated sludge (WAS) rather than primary sludge. Retention times for the sonic treatment are short, on the order of a few seconds.

Reports of the effects of ultrasonic treatment on biogas production are generally favourable, with increases in the production rate reported as in the range of 45-50% (Bartholomew, 2002). A case study of ultrasonic sludge treatment in the United Kingdom, presented in Section 6.0, also reported increases on biogas production of 22%. The biogas production data conflicted with the overall volatile solids reduction, and alkalinity and volatile fatty acids concentration data in the digester, which showed little change as a result of the ultrasound treatment (Kaye, 2007b).

Most of the installations found in the literature are located in Europe, particularly in Germany. Other installations are in Sweden, England and in New Zealand. There have also been a number of full-scale trials including Avonmouth, Wessex, UK and Orange County Sanitation District (Hogan et al., 2004).

The power consumption of an ultrasonic unit is high at about 3.7 kWh per m^3 of wastewater processed (Nemw, 2001). Retrofitting sludge processing facilities for ultrasonic treatment is not difficult. While overall capital costs for the acoustic horns are lower compared to other sludge treatment processes, but because they treat a reduced volume of waste secondary sludge, unit costs per volume treated are higher.

CROWN® Disintegration Process

The CROWN® process, manufactured by the Biogest AG company of Germany, is based on the cavitation phenomenon, in which dissolved gas bubbles in the WAS collapse due to sudden rapid pressure reduction. The collapse causes in the micro-environment a huge increase in temperature and pressure, resulting in a release of mechanical energy as a shock wave to the surrounding matter. The shock wave causes the disintegration of the WAS cells. In the CROWN® process, the WAS is first macerated, and then increased in pressure to 12 bar. The actual "disintegrator" in the process is where the pressure is released and cavitation takes place. Following the disintegrator, the treated sludge is transferred to a "relaxation" tank, prior to further downstream processes such as anaerobic digestion.

Reported benefits of the CROWN® process include:

- Prevention of foaming in anaerobic digestion tanks
- Increase in biogas yield
- Decrease in sludge quantity for disposal
- Enhancement of dewatering
- Increase in electrical power production (presumably from on-site generation).

The process can be supplied as a skid-mounted structure. The manufacturer suggests that only 30-40% of the sludge flow needs to be treated by the CROWN® process to optimize energy use. Several installations in Germany have been reported. A summary of the results for full-scale test installations is provided below in Table 4-5. The test period was typically 12 months. No cost data were reported.

Table 4-5. CROWN® Disintegration Process Installations.

Parameter	Treatment Facility			
	Taunusstein-Bleidenstadt	Ingelheim	Nierstein-Oppenhem	Münchwilen
Trial Date installed	Oct. 2002	Dec. 2002	May 2004	Mar. 2005
P.E. treated	30,000	200,000	20,000	30,000
Sludge Type	6% WAS	3-5% WAS	4% mixed Prim./WAS	6% WAS
Sludge flow rate (m3/h)	6	7	4	4
Energy demand (kWh/m3)	1.42	0.94	1.85	1.85
Improved biogas yield (%)	28	34	18	34
Dig. Sludge volume reduction (%)	22	n.r.	n.r.	n.r.
Improved sludge dewaterability (%)	n.r.	n.r.	n.r.	14.7
Reference	Biogest 2006a	Biogest 2006b	Biogest 2006c	Biogest 2006d

Lysate-Thickening Centrifuge (Baker Process)

Mechanical cell lysis is a process that beaks the cell walls of the waste activated sludge (WAS) biomass, releasing the soluble components to be used as substrate for anaerobic bacteria. Positive results associated with the cell lysis include higher volatile solids destruction in the anaerobic digesters, a higher production rate of biogas, and improved dewatering of the digested sludge.

The process makes use of centrifugal force applied to waste activated sludge to cause cell breakage or disintegration. After thickening in a centrifuge, the WAS is discharged to the patented "lysate ring", in which the solids break down due to collision with each other, due to a high reduction in velocity, and due to shear in passing from the centrifuge bowl to the lysate ring (STOWA, 2006b). Centrate quality is unaffected because the disintegration occurs in the lysate ring after the thickened WAS leaves the centrifuge bowl.

According to the manufacturer's (Lysatec GmbH, Germany) literature, the process resulted in an increase in biogas yield of up to 25% at the Koln-RodenkirchenWWTP, and a volatile solids content of the dewatered sludge of less than 50% (STOWA, 2006b). Although energy consumption is anticipated to increase by approximately 20% with adoption of the lysate ring, overall savings of 40,000 €/yr are estimated for a treatment plant capacity of 100,000 p.e.

4.4.1.2 Thermal Energy Recovery Only

Boilers

Boilers have been used for heat recovery from digester gas for many years. They are one of the least capital-intensive energy recovery technologies available. Boilers require little gas pretreatment and have an energy recovery efficiency of 75 – 80%. In boilers, digester gas is combusted to produce steam or hot water. Boilers do not generate electricity or mechanical power. Under summer conditions, they may not be able to use all the digester gas produced, resulting in flaring (wasting of the excess gas).

Incineration

Incineration, also called thermal oxidation, is the process of combusting organic waste at high temperature in the presence of oxygen (air). This process involves five main steps: 1) Dewatering, 2) Drying, 3) Combustion, 4) Air pollution control equipment and 5) Ash management.

Dewatering is used to increase the solids concentration of the feed to between 15 and 35%. Drying raises the sludge temperature to the point that water in the solids evaporates. Combustion is used to destroy the volatile fraction in the sludge. This is an exothermic reaction that results in the production of inert ash and hot gases. Thermal energy can be recovered from the hot gases.

Two types of incineration technologies are used for sludge combustion: multiple hearth furnaces (MHFs) are an older technology still in use in some municipalities, and the newer technology fluidized bed furnaces (FBFs).

Incineration is a commercially proven technology. The U.S. EPA (2003) reported that in 1993, 343 biosolids incinerators were in operation in the United States. Of these, approximately 80% were MHFs and 20% were FBFs. Use of FBFs technology for wastewater solids incineration has considerably increased the last 10 years because they are more efficient, more stable and easier to operate than MHFs (U.S. EPA, 2003). Several plants in North America have replaced their MHF installation by a FBF installation (Dangtran et al. 2002). Hand-Smith (1999) reported that several incineration plants were operating in the UK in 1999 and others were in their commissioning phase.

Thermylis® HTFB is a High Temperature Fluidized Bed Incineration Technology developed and commercialized in France. This process operates with a temperature around 840°C. About 14 installations have been constructed from 1995 to 2007 in North America and France. Three new installations are planed in Ontario, Canada in 2008, each with a capacity of 110 dry MT/d.

In Leeds, England, sludge is dewatered prior to injection of the sludge cake in a fluidized sand bed furnace at a temperature higher than 850°C. The bed is fluidized by pre-heated air, which evaporates the remaining water and incinerates the sludge to an inert ash. A heat recovery system recovers heat from the flue gases to pre-heat the combustion air, and to generate steam, which is then used to pre-dry the feed sludge and generate electricity in a steam turbine. Mercury is removed from the flue gases by a series of sophisticated processes including an adsorption stage.

Incineration processes tend to be complex because of the technologies used for thermal energy recovery, fly ash capture, and the need in many cases to remove metals such as mercury. If the solids have been dewatered sufficiently, then the process will be autothermal; if too much moisture remains in the feed sludge, however, supplementary fossil fuel will be required. Because of the process complexity, the incineration processes have a relatively large land footprint. Bottom ash and fly ash produced during the combustion can be used as additive for cement and brick manufacturing (see above). Ash may also be used for phosphorus recovery (see above); otherwise the ash can be landfilled.

In the Netherlands, a novel use of off-site surplus heat from a solid waste incinerator to maintain year-round temperature control in activated sludge tanks is presented as a Case Study in Section 6.0. Because the secondary treatment process can operate at higher temperatures, a plant upgrade for nutrient removal could make use of smaller tank design

4.4.1.3 Combined Heat and Power from Digester Gas (as Electricity)

Internal Combustion Engine-Generator Sets (Cogeneration)

The use of digester gas in internal combustion engines to generate electricity in an engine/generator set is called cogeneration. The technology has been used with digester gas for many years. The thermal energy recovery efficiency of cogeneration is reported to be in the range 45-50%, while the electrical energy recovery efficiency is 30-5%. Thus the overall energy recovery efficiency is reported to be 75-85%. Contaminants in the digester gas that must be reduced for cogeneration include moisture, hydrogen sulfide and siloxanes. In the United States, use of cogeneration may be limited due air quality regulations which restrict emissions of nitrogen oxides (NO_x) a product of gas combustion.

Turbines

Microturbines are a newer energy recovery technology suitable for small to mid-size plants. Microturbine units are available in sizes in the 50-150 kW range. In microturbine operation, combustion of biogas drives a turbine fan blade on a shaft, which rotates through the generator to produce variable frequency three-phase alternating current (AC) power. The electrical efficiency of micro-turbines is approximately 26-27% at 30 kW operations. With combined heat and electric power recovery, the overall efficiency is between 70-90%. Pretreatment to remove water and siloxanes is recommended. Microturbines are growing in popularity in areas of the U.S. impacted by air quality regulation, because the produce low NO_x emissions during combustion of the digester gas.

For larger wastewater treatment plants, larger simple cycle or combined cycle gas turbines may be more relevant. In simple cycle turbines, atmospheric air is compressed to high pressure, with a portion of the compressed air diverted through digester gas burners to raise the temperature of the compressed air. This very hot gas is mixed with the rest of the compressed air and directed to the power turbine. Expansion of the hot compressed air past the turbine blades rotates the shaft. The electrical energy efficiency is typically in the 30-35% range. (Virginia DEQ, 2007).

Combined cycle turbines operate similarly to the simple cycle turbines, but the hot exhaust gases from the turbine, instead of being vented directly to atmosphere, are directed to a waste heat boiler to generate steam. This steam drives a steam turbine generator to make additional electricity, resulting in an increase in the cycle electrical efficiency to 50 percent or more. A combined cycle turbine is more capita intensive than a simple cycle turbine because of additional capital costs for a waste heat boiler, steam turbine, and cooling system. The operating cost per unit of electricity produced is lower for a combined cycle turbine compared to a simple cycle turbine due to the improved energy recovery (Virginia DEQ, 2007).

Fuel Cells

Fuel cells produce electrical power directly through an electrochemical reaction using hydrogen and oxygen. The hydrogen is produced from digester gas while air provides the oxygen. First, fuel gas is combined with water in the cell where the gas is reformed to hydrogen and fed to the fuel cell anode. Air is fed to the cathode. The process of hydrogen moving from the anode to the cathode through an electrolyte creates electricity.

There are several types of commercially available fuel cells: low temperature (includes Phosphoric Acid, Proton Exchange Membrane and Alkaline types) and high temperature (Molten carbonate and solid oxide types). Overall fuel cell efficiency varies by type between 47 – 87%. Only the phosphoric acid and molten carbonate fuel cells have been investigated at full-scale to any extent. Fuel cell capacities at installations in North America (U.S.) range from 200 kW to 1 MW. Demonstration projects have been conducted in King County (WA), Los Angeles and Las Virgenes (CA), Portland (OR) and New York City. Capital costs for fuel cells currently are very high compared to other technologies, although the operating costs can be very low on the order of \$0.01/kWh. Fuel cells have very low emission rates of NO_x and SO_x, and are well-suited to locations that are impacted by stringent air quality regulations.

4.4.1.4 Combined Heat and Power from Digester Gas (as Mechanical Energy)

Direct Drive Engines

The energy in digester gas can also be used as recoverable mechanical energy. The theoretical benefit of direct drive units is that they convert gas energy directly into mechanical energy without the intermediate step of conversion to electricity. The primary direct application is to drive aeration blowers and pumps. The reported recoverable energy conversion efficiency for direct drive applications is 46% (thermal) and 26-28% (electric). Direct drive engines have been in use for many years. A 1981 study by Battelle Pacific Northwest Labs suggested that 50-80% of the mechanical energy for wastewater treatment could be provided by digester gas. As with cogeneration units, the digester gas may need to be treated for moisture, hydrogen sulfide and siloxanes prior to combustion in the engine. Emissions of NO_x may be an issue for treatment facilities affected by stringent air quality regulations.

Stirling Engines

The Stirling engine differs from an internal combustion engine; it is a closed-cycle piston heat engine where a working gas (typically helium or hydrogen) is permanently contained within the cylinder. The Stirling engine uses the potential energy difference between a heat source and a heat sink to establish a cycle of expansion and contraction of the working gas within the engine, thus converting a temperature difference across the machine into mechanical power. The greater the temperature difference, the greater the power produced.

Stirling engines can operate on digester gas burned in an external combustion process to create the temperature difference. The change in working gas temperature will cause a corresponding change in gas pressure, while the motion of the piston causes the gas to be alternately expanded and compressed in a sealed chamber. When the gas is heated in the sealed chamber, the pressure rises and acts on the piston to produce a power stroke. When the gas is cooled, usually by water, the pressure drops and the piston compresses the gas on the return stroke, thus yielding a net power output.

Because the working gas is separate from the heat source, a wide range of heat sources can be used, including digester gas, solar or waste heat. Since the combustion products do not contact the internal moving parts of the engine, a Stirling engine can run on digester gas containing siloxanes or hydrogen sulfide without damage or without costly pretreatment.

Stirling engines have emerged as an alternative energy source when heated by the combustion of landfill gas. At least one firm has a commercial 55kW unit which claims a 30% electrical efficiency and an 80% combined heat and power efficiency. The Corvallis treatment plant in Oregon uses a Stirling engine to generate onsite power.

4.4.2 Sludge-to-Syngas Processes

Sewage sludge can be converted in energy-rich gases that can then be used to generate electricity. The technologies developed to perform such treatment are based on pyrolysis and gasification.

4.4.2.1 Gasification

Gasification is a process that takes place in two steps. In the first step, the volatile fraction of the solids is transformed, in absence of air (pyrolysis), into a carbon-rich substance called "char". This transformation occurs at a temperature around 600°C or less. In the second step, the char is gasified in the presence of oxygen or air. This reaction produces a gas called syngas. The latter typically contains nitrogen (N_2), carbon dioxide (CO_2) and carbon monoxide (CO) and hydrogen (H_2). Syngas can be used as fuel to generate electricity and heat. Most of the full-scale gasification installations are operating in Europe, and mainly in Germany.

Examples of gasification technologies are the KOPF Gasification Technology (Germany) and the EBARA fluidized bed gasification technology (Japan).

KOPF Process

The main components of the KOPF gasification technology are; a solar drying unit, a fluidized-bed gasification unit, a gas engine unit for energy recovery and a post combustion chamber for burning excess syngas. A unique feature of the process is the solar drying unit, which dries the wet digested sludge to a solid content of between 70 and 85% over a period of 2 to 8 weeks, depending on the weather conditions. Since this thermal energy is free, the energy and operating cost requirements compared to other processes using fossil fuel for drying are substantially lower, with a reduced carbon footprint. With 36 sludge dryers operating in Europe, solar drying appears to be completely adaptable to the European climate. The time required for drying will be longer in winter than in summer.

In the gasification reactor, operating at 900°C, pre-heated air is used to ensure the fluidization of the bed. Inside the reactor, dried solids are converted into inert granules and combustible gas. The gas is recovered and cooled to a temperature below 150°C and dried. Natural gas is required for plant start-up, but after the start-up phase, no external fuel is needed. The gas engine produces electricity, which is used to operate the gasification process and to offset the energy demand of the sewage works. Recovered thermal energy is used to heat the digesters.

EBARA Process

There is very limited information about the full-scale plants using the EBARA technology. The technology co-treats municipal sludge with other solid wastes, including fly ash, waste plastics, liquid waste, medical waste and municipal solid waste. Because of the mixed nature of the feed, many by-products can be recovered by the EBARA process, including ferrous and non-ferrous metals, a glass aggregate, and metal recovery from boiler and filter ashes. As of spring 2002, six commercial TwinRec units were in operation (Steiner et al. 2002).

4.4.3 Sludge-to-Oil Processes

Sewage sludge can be converted to oil, which then can be used as an industrial fuel in boilers to produce steam, or it can be combusted in an engine to produce electricity (Bridle, 2004). The oil can also be used for biodiesel production (Salter, 2006; Dufreche et al., 2007;

Liang et al., 2007). The processes that can be used to produce oil from sludge include pyrolysis and hydrothermal treatments.

Pyrolysis is the conversion of waste in the absence of air to char and syngas. The syngas may be converted to oil depending on the reactor operating conditions. There are two categories of pyrolysis; slow pyrolysis and fast pyrolysis. Slow pyrolysis does not produce oil. This process operates at temperatures below 315°C with a long vapor residence time in the reaction zone.

Fast pyrolysis produces oil as a product. It takes place between 425 and 538°C, and has a shorter vapor residence time than slow pyrolysis. In fast pyrolysis, the sludge is dried to less than 10% water to minimize the quantity of water in the final oil stream. The vapor residence time is less than 2 seconds, with rapid cooling and condensation of the vapor to produce the bio-oil.

Commercial pyrolysis technologies for sewage sludge treatment are EnerSludge™ Technology (Bridle, 2004; Bridle and Skrypski-Mantele, 2004) and SlurryCarb™ Technology (EnerTech, 2006).

EnerSludge™

EnerSludge™ has been developed over the past 15 years, with extensive pilot plant demonstration programs conducted in Australia and Canada (Bridle et al. 1989; Gough et al. 1991). The first commercial demonstration installation was constructed at the Subiaco MWTP in Perth, Western of Australia. Although the process requires liquefied petroleum gas (LPG) continuously, it produces net energy in the form of oil. Overall 45% of the energy in the biosolids is converted to oil (Bridle, 2004). To increase efficiency, the oil could be combusted in a diesel engine to produce electricity and recoverable heat. After a 16 month trial period, the process was shut down and replace by a lime amended process because it was not considered cost-effective (GVRD, 2005).

SlurryCarb™

SlurryCarb™ was developed in the United State. The technology operates in 7 steps including, sludge preparation, slurry pressurization, slurry heating at a temperature between 270°C and 325°C, reaction, dewatering/drying, filtrate processing and recycle and fuel utilization. The process appears to be a type of slow pyrolysis.

SlurryCarb™ converts sludge into a fuel called E-fuel and CO_2 gas. The E-fuel is the carbonized product (char) of the sludge. The product can be recovered either as slurried E-fuel or as Dried E-fuel. These two materials can serve as fuel in cement kilns (EnerTech, 2006).

A SlurryCarb™ pilot plant operated in the city of Ube in Japan from 1997 to 2000 as demonstration project. A new full-scale installation is now being constructed at the Rialto MWTP in California, with an anticipated start-up early in 2008. The facility has been designed to treat 687 MT per day of sludge, at 10-31% dry solids, recovering approximately 120 MT of dried pellets per day that can be used as renewable fuel in a near by cement plant.

4.4.4 Sludge-to-Liquid Processes

4.4.4.1 Supercritical Water Oxidation

Supercritical water oxidation (SCWO) is a process that occurs in a water phase at temperatures and pressures above the super critical point of water. The process has been under development for more than twenty years (Svanström et al., 2004). Both raw and digested sludges without thickening or dewatering, with dry solids concentrations as low as 3% (Levlin, 2004). Operation takes place at a temperature around 374°C, a pressure higher than 22,000 kPA (220 bars). Pure oxygen used in the process is usually produced using cryogenic air separation (Bernstein, 1999), which requires energy for operation.

This process is capable of converting more than 99.99% of the chemical oxygen demand (COD) present in the sludge to carbon dioxide within 1 minute (Stendahl and Jäfverström, 2004). The effluent from the SCWO reactor is a slurry of inorganic ash in a water phase. Inorganic components like phosphates (P_2O_5 content of 6.3-18.4 %) and/or coagulants can be recovered from the ash (Stendahl and Jäfverström 2004).

During sludge treatment with SCWO, energy can be recovered directly by heat exchange in the reactor, or from the effluent leaving the reactor. The heat in the effluent is transferred to a water stream through a heat exchanger unit. An energy balance indicates that the thermal energy recovered as hot water is greater than the energy inputs.

Advantages of SCWO compared to incineration are: 1) no need to dewater the sludge at very high dry matter concentration and 2) low operating temperature. The main drawback of SCWO is that only heat can be recovered. This may be a limitation of the process because while wastewater treatment plants need both electricity and heat to operate, electricity is the principal energy need.

Examples of SCWO technologies are Aqua Reci® Technology (Stendahl and Jäfverström, 2004), Aqua Citrox® Technology (Gidner and Stenmark, 2001) and the Athos® Technology (Veolia, 2007). Only the last technology has been developed and implemented at full-scale, with five operations in France (3), Belgium (1), and Italy (1). Detailed operating or financial information regarding those plants is not available.

The SCWO technology is complex, requiring significant capital expense and land. Cryogenic oxygen equipment is required further adding to the complexity, expense and carbon footprint.

4.5 Market Drivers

Market drivers can be defined as the key factors that incite manufacturers to develop and promote a specific product, service or technology that may fulfill the needs of the market. In this report, at least four main drivers in the area of energy and resource recovery from sludge have been identified. The four drivers are sustainability/environmental concerns, energy cost and type, resource quality and quantity and regulation/legislation.

4.5.1 Sustainability/Environmental Concerns

Sustainability and environmental concerns with respect to sludge disposal involve at least three issues, namely soil pollution; greenhouse gas emissions, and resources depletion. These three factors force the development of more environmentally friendly technologies.

Direct use of sewage sludge in agriculture is certainly an economic option. There is however, a broadly-supported perspective in some countries that direct agricultural application of sewage sludge is not a sustainable approach. Such concerns contribute to the exploration and development of new paths for sludge management. Indeed, elevated concentrations of metals in the sludge may result in a prohibition of land application, resulting in the sludge being used in other applications, such as a supplementary fuel in cement kilns.

The contribution of wastewater treatment to global warming is now of major interest. Energy produced from sludge is considered as renewable energy that can be directly used for the wastewater treatment, reducing the facility's dependency on conventional electricity. The greater the quantity of energy which the wastewater treatment industry can produce, the more this industry can help to reduce emissions of greenhouse gases (GHG). Thus technologies that can be used to recover as much energy as possible from sludge are viewed as environmentally friendly technologies. This positive ecological impression enables the process manufacturers to position themselves well in the market.

A resource such as phosphorus is now considered as a non-renewable resource. The world stock of remaining P rock may no longer be available within 50 to 150 years. Currently, there is no obvious replacement alternative except the recovery of P from wastewater sludge, and other organic wastes such as livestock manures or dairy wastes. As long as no alternative to P rock becomes available, research and development for efficient and cost-effective recovery of P from sludge will be a major area of technological development.

Sustainability and environmental concerns are highly influenced by public perception. Thus the success of any technology will strongly depend on how this technology is perceived by the public. This topic is discussed in detail in the section IV.4 under Social Feasibility.

4.5.2 Energy Cost and Type

Energy (mainly electricity) is a key requirement for the operation of wastewater treatment plants. Energy consumption represents about 25% of the O&M costs of MWTPs (Smith and Clark, 1995).

It is critical for wastewater treatment plants to choose the most efficient and cost-effective energy strategy, because an increase in energy cost directly results in an increase in the O&M costs of the plants. A typical example is the situation of the UK where the energy cost has doubled the last three years (Kaye, 2007). In order to cope with such inflationary situations, on-site generation of energy from sludge has becomes an attractive option, not only in term of energy cost savings, but with the added benefit of reducing greenhouse gas (GHG) emissions due to fossil fuel use at electricity generating stations, as already discussed above.

The results of the literature survey presented earlier in this chapter indicate that there are a large number of technologies capable to recover energy from sludge. It is clear, however, that none of them is efficient enough to fully extract all the energy available in wastewater. In the early chapters of this report, it was noted that experimental laboratory research showed that sewage contains at least 10 times the energy required to treat it. This means that new technological developments and/or the optimization of the current technologies are necessary to fully take advantage of the maximum energy available in sewage and sludge. As energy costs continue to rise, this will become a major goal for wastewater treatment plants.

Successful technologies will be those offering the highest net energy output. If excess energy is generated, this excess can be employed to run the other processes that may be used to recover other resources from sludge such as nutrient and building materials.

The type of energy produced may also be an important aspect of the success of the technologies. Wastewater treatment plants need both electricity and heat. Some of the technologies currently available such SCWO produce only heat. Effort should probably be oriented towards the development of technologies capable to generate electricity and heat. Such technologies may have a broader market niche.

4.5.3 Resource Quality and Quantity

The quality of the end product from sludge processing is an issue that affects the development of new and more efficient technologies. If high quality product is required, more efficient technology is then necessary. A typical case is phosphorus. Different grades or forms of phosphorus can be recovered from sludge depending on the technology used. But the products (e.g. iron phosphate, calcium phosphate, phosphoric acid, etc.) recovered do not have the same market value. Potentially valuable phosphorus products are phosphoric acid and calcium phosphate (Stark, 2005). Calcium phosphate is the raw material used by industry to manufacture phosphoric acid.

Even if a technology is capable to produce calcium phosphate, attention should be paid to the impurities the product may contain, such as heavy metals. The Dutch experience of phosphorus recovery for the phosphate industry has shown that low quality product will not readily find an outlet (Roeleveld et al., 2004).

The quantity of product that can be recovered is also a factor that is considered when decision makers decide the technologies they wish to pursue. The wider the market for the potential product, and the ability to respond to that market, the more attractive the technology will be. This is particularly important when the product is for commercial sale, because greater production will potentially generate more revenue. Higher revenues reduce the pay back period of the technology.

4.5.4 Regulation/Legislation

Regulation and legislation can be used as stimulating factors for the development of technologies in the area of energy and resource recovery. This is typically true in a context where there is no direct economic driver. Sweden is an example of country where regulation has been used to initiate the development of technologies of phosphorus recovery from sewage sludge. As indicated in Chapter 3.0, this country has established regulations requiring wastewater treatment plants to recover nutrients such as phosphorus in case of incineration (Hultman et al., 2001). In the Netherlands, a ban on land application of sludge has resulted in wider use of the sludge as a fuel in cement kilns, for example.

Other types of regulations can help to determine the advancement of different technologies. In the U.S., for example, air regulations governing ground level concentrations of ozone and NOx favor the adoption of energy recovery equipment such as microturbines and fuel cells which produce low emissions of the air pollutants.

4.6 Feasibility of Energy and Resource Recovery

Feasibility of energy and resource recovery basically depends on the following three factors, technical feasibility, economic feasibility and social feasibility. Each of the three factors is discussed below.

4.6.1 Technical Feasibility

It is technically feasible to recover energy and resources from sludge by anaerobic digestion, incineration, pyrolysis etc. Based on the results of the literature survey presented earlier in this chapter several processes for energy recovery are available for the wastewater industry. Table 4-6 summarizes the energy input and output of the main established energy recovery technologies. Technologies for which data were available in the literature indicate that their energy output is higher than the amount of energy required for operating the processes. An energy output up to 16 times the amount consumed can be achieved by some of the technologies, such as EnerSludge™ and Cambi®.

It has also been established that resources such phosphorus can be recovered by extraction-precipitation with a recovered efficiency up to 60-70% or even higher (Stark, 2005; Levlin et al., 2004 Stendahl and Jäfverström, 2004). Although P recovery on full-scale is a feasible option, this practice is in its early stages. It should also be recognized that iron and aluminum salts used to remove P from wastewater limit the possibility of recovering phosphorus as calcium phosphate or as struvite. Calcium phosphate is the raw material needed by the phosphate industry. There is currently no commercial method to produce phosphate from iron phosphate (Stark, 2005). Thus processes using biological P removal in MWPTs should be prioritized in a first approach. Only when those processes are not capable of meeting the requirements for effluent P concentration, that chemical process should be considered.

Table 4-6. Summary of energy input and output of the main established energy recovery technologies.

Process Category	Technology	Total Energy Input	Total Energy Output
	Type	kWh/dry MT	kWh/dry MT
Sludge-to-Biogas			
Thermal hydrolysis	Cambi®	0.3	6.0
Thermal hydrolysis	BioThelys®	N.A.	N.A.
Cell destruction	MicroSludge™	502	1358
Cell destruction	Ultrasonic	141	N.A.
Sludge-to-Syngas			
Gasification	Kopf	100	1400
Gasification	EBARA	N.A.	N.A.
Incineration	Thermylis® HTFB	N.A.	N.A.
Sludge-to-Oil			
Pyrolysis	EnerSludge™	120	1966
Pyrolysis	SlurryCarb™	712	758
Sludge-to-Liquid			
SCWO	Athos®	900-1200	1,680

N.A. indicates information not available.

4.6.2 Economic Feasibility

Cost is a decisive aspect in energy and resource recovery from sludge. Two types of costs are associated with each technology: The capital cost and the operating and maintenance (O&M) costs. If the present worth cost (Capital and O&M) of a technology which looks environmentally attractive is not affordable, the technology is unlikely to be adopted unless other market drivers come into effect.

Determination of the economic feasibility of energy and resource recovery from sludge is a complex issue. For each technology, this depends on several factors. In general, the more complex the technologies are, the more costly they are. Capital and O&M costs depend on the type of technology, the size of the installation, the type and number of input material for the operation of the installation, plus local conditions such as land and labor costs. Economic feasibility will also depend on the type of resource that is to be recovered. The cost may depend as well on the efficiency or product quality that must be achieved. Higher efficiency or quality requires higher capital and O&M costs (Hultman et al., 2003).

In order to roughly compare the technologies, a common metric can be used to relate the cost of different established technologies reviewed in the previous sections of this chapter. Such a unifying factor can be defined by dividing the cost by the dry MT of sludge processed per year. Table 4-.7 shows a large variation between the costs of the energy recovery technologies that are currently marketed. The costs are from different years, but the years are close enough (difference is less than 10 years) to provide a basis for comparison.

Table 4-7. Summary of capital cost of established energy recovery technologies from sludge.

Process Category	Technology	Capital Cost	Capital Cost	O&M Costs	O&M Costs
	Type	*US$/dry MT/yr*	*Data Year*	*US$/dry MT*	*Data Year*
Sludge-to-Biogas					
Thermal hydrolysis	Cambi®	650-1700	2005	396	2000
Thermal hydrolysis	BioThelys®	N.A.	*Not Applicable*	N.A.	*Not Applicable*
Cell destruction	MicroSludge^{TM}	488	2006	75-131	2006
Cell destruction	Ultrasonic	533-842	2006	20-25	2006
Sludge-to-Syngas					
Gasification	Kopf	N.A.	*Not Applicable*	N.A.	*Not Applicable*
Gasification	EBARA	N.A.	*Not Applicable*	N.A.	*Not Applicable*
Incineration	Thermylis$^{\circledR}$ HTFB	N.A.	*Not Applicable*	N.A.	*Not Applicable*
Sludge-to-Oil					
Pyrolysis	EnerSludge^{TM}	1640-1954	2007	86-113	2007
Pyrolysis	SlurryCarb^{TM}	1063-1772	2006	101-104	2006
Sludge-to-Liquid					
SCWO	Athos®	N.A.	*Not Applicable*	N.A.	*Not Applicable*

N.A. indicates information not available.

The calculated estimates based on literature, such as the costs presented in Table 4-7, are certainly helpful, but alone are insufficient to accurately define the cost of a technology. A more detailed assessment involving site-specific circumstances must be completed in order to better understand the level of expenses required and possible savings.

To be attractive, technologies for energy and resource recovery must be cost-effective. This is currently not always the case. Many projects generally fail because of the high capital and O&M costs. Examples of such projects are phosphorus recovery (Hultman et al., 2003; Roeleved et al., 2004) and building material production from sewage sludge (Okuno et al., 2004). The main issue with resource recovery is related to manufacturing cost of the products versus the market price.

Currently the economic feasibility of phosphorus recovery from sewage sludge is unrealistic mainly because the costs of the technologies are substantially higher than the cost of mined phosphate rock (Hultman et al, 2003; Roeleved et al .2004). It has been observed in Sweden that the cost for recovering P is two times higher than the market price (Hultman et al., 2003). Technologies using many chemicals as input may be 3 to 8 times expensive than the world price of raw phosphate (Schaum et al., 2007; Balmér, 2004). A survey conducted by Jeanmaire (2001) a few years ago revealed as well that the cost for P recovery is unrealistic when compared with the cost of P rock used in the industry, and the cost of phosphorus used as fertilizer (Table 4-8).

Table 4-8. Cost comparison of recycled and rock P and Fertilizers (Adapted after Jeanmaire, 2001).

Use of P	Optimized costs of recovered P €/MT of P (US$/MT of P)	Costs of P rocks in North Europe €/MT of P (US$/MT of P)
Recycling in the industry	10,200 (15,200)	320 (480)
Use of P	**Cost of recovered P as fertilizer €/MT of P (US$/MT of P)**	**Costs of fertilizer €/MT of P (US$/MT of P)**
Recycling as fertilizer	1,235-2,833 (1,853-4,250)	850-1,153 (1,275-1,730)

Since global phosphorus mineral reserves are limited and phosphorus is not an endless resource, it is likely in the future that the difference between the cost of recovered P and the cost of mined rock P will decrease as the quantity of rock available further decreases. By today's prices, however, it remains a challenge to develop P recovery technologies that are cost effective.

Similarly, the Japanese thermal solidification experience for building material production showed that thermal solidification processes from sludge incineration ash are technically feasible, but many of them are not economical because the manufacturing cost is higher than the market price (Okuno et al., 2004). Thermal solidification processes are usually energy intensive processes. To make the extraction of building materials competitive energy efficient technologies are needed.

4.6.3 Social Feasibility

Technology that can not meet social acceptance will usually be difficult to market. Social acceptance of a technology depends on the values adopted by a nation's citizens. For example, although sludge incineration is practiced in North America, it is often not well accepted by the public. This is different from European countries like Germany (ED, 2002b) or Switzerland (ED, 2002 a) where sludge incineration seems to be well accepted. In Netherlands as well, almost half (48%) of the quantity of sewage sludge produced is incinerated (Uijterlinde, 2007).

Social acceptance of a technology depends also on the types of inputs used. Some of the technologies described earlier involve use hazardous chemicals (e.g. H_2SO_4, NaOH). In the next Chapter of the report focusing on emerging technologies, many technologies, particularly those used to recover phosphorus, use many chemicals such H_2SO_4, NaOH, $Ca(OH)_2$, hydrogen peroxide, and others.

Technologies based on chemical use are more susceptible to be rejected the public if the chemicals are potentially hazardous to human health and the environment. For example, use of H_2SO_4 in resource recovery may be regarded negatively because it is potentially harmful. It is highly corrosive and can severely burn the skin and eyes of exposed individuals. With regard to the environment, this chemical can burn any plants, birds or animals exposed to it due to accidental releases. It exhibits chronic toxicity to aquatic organisms. There are therefore potential human and ecological risks associated with the production, transportation, storage and use of such chemical.

From a life cycle perspective, the use of chemicals implies also that they have to be produced somewhere, transported from the production site to the manufacturing plant site and then stored on-site. Both energy use and emissions are associated with the production and transportation of all chemicals. Processes with several separate steps are generally viewed as complex processes, land consuming, high capital and O&M costs. This is typically the case for processes such as gasification

Technologies which result in release of pollutants will have less chance to be accepted by the public. In general biological processes using microorganisms, present in nature, to carry out treatment are perceived positively by public as a natural processes. This is for instance the case of anaerobic treatment processes. However, issues such as odour may create public opposition.

Lastly, another factor related to social acceptance of a technology is the status of the technology. Processes that have been established in different countries with multiple installations are more likely to be viewed positively by the public than those that are in development or untested. One example of an accepted technology is the Cambi® process.

5.0 Future Developments/Emerging Technologies

The technologies described in this chapter for resource and energy recovery have not been commercially established, compared to those in the previous chapter. This section therefore addresses those processes which may be considered as emerging technologies.

5.1 Emerging Technologies for Resource Recovery

5.1.1 Emerging Technologies for Phosphorus Recovery

Several technologies are currently under development for phosphorus recovery either from raw sewage sludge or sludge incinerator ash. Table 5-1 summarizes the name of the technologies, their origins, the final product and the inventors. Most of the technologies originate from Europe particularly from Germany, Sweden and Denmark. The feasibility of recovering phosphorus from sludge incinerator ash is under investigation in the Netherlands; this is presented as a Case Study in Section 6.0.

The emerging technologies for phosphorus recovery are mainly based on physical-chemical and thermal treatment to dissolve phosphorus. Phosphorus is then separated from heavy metals by precipitation. As already indicated in a previous chapter, depending on the technology used, phosphorus can be recovered from sludge as iron phosphate, calcium phosphate, phosphoric acid and struvite (magnesium ammonium phosphate, MAP).

Table 5-1. **Emerging technologies for phosphorus recovery from sewage sludge.**

Technology	Origin	Needs	Final Product	Inventor
Recovery from Sludge				
KREPO	Sweden	Heat, H_2SO_4, NaOH, Fe	Iron (III) P	Kemira K.
Seaborne	Germany	H_2SO_4, NaOH, $Mg(OH)_2$	MAP	Seaborne E.R.L
Aqua-Reci®	Sweden	Heat, HCl, O_2	Iron or Calcium P	Chematur / Feralco
Kemicond™	Sweden	H_2SO_4, H_2O_2, Polymer	Ferric P	Kemira K.
Recovery from Ash				
BioCon	Denmark	H_2SO_4, Ion exchanger	Phosphoric acid	PM Energi A/S
SEPHOS	Germany	H_2SO_4, NaOH, Ca^{2+}	Aluminium and Calcium P	Institut WAR
SUSAN	Europe	Heat	Phosphorus, heavy metals	

5.1.1.1 KREPO Technology

KREPO is an acronym for Kemwater REcycling PROcess (Hultman, 1999; Hultman, et al. 2001). This technology developed in Sweden uses heat, pressure and sulfuric acid to dissolve phosphates, metals and a large fraction of organic compounds from thickened sludge.

The raw sludge is initially thickened to 5% dry solids in a centrifuge. The main steps involved in this technology are: initial acidification, heating with steam, hydrolysis in a pressurized reactor, organic sludge separation and precipitation of iron phosphate from the centrate.

A full-scale KREPO Technology system was operated in 1995 at the Öresundsverket MWTP in Helsingborg Sweden (Hansen et al., 2000). Following the experience of the Öresundsverket MWTP, the city of Malmö planned a full-scale KREPO plant (SCOPE, 2001), although the plan was later abandoned (Hultman et al., 2003). Reasons for the decision were not discovered.

A more detailed description of the process is found in Appendix C.

5.1.1.2 Kemicond^{TM} Technology

Kemicond is an acronym for Kemira sludge conditioning. This technology was developed in 2003 by modifying the KREPO Technology (Berg and Shaum, 2005) described above. The process which treats raw sewage sludge, consists of a acidification with sulfuric acid and hydrogen peroxide (H_2O_2), followed by a two stage dewatering unit (Cornel et al., 2005). During the chemical treatment, metals such as iron phosphate and hydroxides are dissolved initially by the acid. Peroxide addition oxidizes the dissolved iron (II) into iron (III), with the dissolved phosphates re-precipitated as ferric phosphate for recovery by solids dewatering. Polymer is used to enhance the dry solids recovery.

The first plant with Kemicond^{TM} technology started up in Stockholm, at the Käppala MWTP. Because the main purpose of the Käppala project was to improve sludge dewatering and reduce sludge transport (Manhem and Palmgrem, 2004), phosphorus recovery was not investigated in detail. In June 2006, Oulun Vesi Water Utility in Oulu Finland, chosen Kemicond for handling sludge from the city's waste water treatment processes. The technology will be installed at the Taskila MWTP which has a capacity of 150,000 P.E (Kemira, 2006a).

The Kemicond^{TM} technology appears considerably less complex than KREPO because there is no pressurized heated reactor. The land and carbon footprint for the Kemicond^{TM} process should be considerably less than its fore-runner.

A more detailed description of the process is found in Appendix C.

5.1.1.3 Seaborne Technology

The Seaborne technology was developed in Germany by the Seaborne Environmental Research Laboratory (Berg and Shaum, 2005). This technology is based on a combination of several processes, including incineration, acid treatment, desulphurization, methane production, heavy metals separation, and struvite precipitation

In the first step, metals and nutrients in dewatered anaerobically digested sludge are dissolved by lowering the pH with an acid solution. The organic residual and the soluble compounds are separated using a centrifuge. The organic residual is incinerated. In the second step, hydrogen sulfide in digester gas is used to precipitate the metals from the centrate. This allows for purification of the digester gas. The metals are separated by filtration. In the third and final step, sodium hydroxide and magnesium oxide are added to the filtrate. This results in the precipitation of struvite (MAP).

The first large-scale Seaborne pilot plant was built between 2005 and 2006 at the Gifhorn wastewater treatment in lower Saxony, Germany. The Gifhorn plant has a capacity of approximately 50,000 P.E. The Seaborne installation has a capacity of 1000 dry MT of solids per year (Müller et al., 2007). The process evaluation is still ongoing.

Reported advantages of this technology are (1) the recovery of multiple nutrients with apparently no heavy metals and organic pollutants and (2) H_2S-free biogas (Müller et al. 2007). The complex technology however requires more unit processes than other technologies.

A more detailed description of the process is found in Appendix C.

5.1.1.4 BioCon

BioCon Technology was developed for recovering phosphorus as phosphoric acid. Ash from sludge incineration is leached with sulfuric acid. The acid solution is then subjected to a number of ion exchange columns, Regeneration of the ion exchange columns yields products including ferric chloride ($FeCl_3$), potassium bisulfate ($KHSO_4$) and phosphoric acid (H_3PO_4) (Hultman, et al. 2003; Levlin et al., 2004).

Although BioCon Technology has not yet been investigated at full-scale, it has been studied at a pilot-scale at a MWTP near Aalborg in Denmark (Hultman et al, 2001). A full-scale plant was anticipated for the city of Falun, Sweden, but the project was abandoned (Hultman et al., 2003). Reasons for the decision were not discovered.

A more detailed description of the process is found in Appendix C.

5.1.1.5 SEPHOS Technology

SEPHOS is an acronym for the SEquential precipitation of PHOSphorus. This technology concept has been initiated in Germany (Berg and Shaum, 2005). The inventors expect that the SEPHOS Technology will produce aluminium phosphate and calcium phosphate using ash from sludge incineration. The first product could be used in the electrochemical phosphate industry while the second could be reused as fertiliser.

In the first stage or sequence, incinerated sludge ash is mixed with sulfuric acid to decrease the pH to 1.5. After separating the solids from the liquid phase, the pH of the liquid is re-adjusted to around 3.5. to precipitate phosphorus and aluminum present in the liquid as AlPO4. No other metals are present at this pH, because copper and zinc only precipitate at pH higher than 3.5 (Schaum et al., 2007). In the second stage recovered aluminum phosphate can be dissolved again by addition of a caustic solution to high pH (12-14). With lime addition, the dissolved phosphorus can be precipitated as calcium phosphate.

No full-scale or pilot scale demonstration of the SEPHOS technology has been conducted yet, so technical operating and economic details are lacking.
The process is relatively simple, requiring only pH adjustment by chemical addition. It requires no heat which reduces the carbon footprint for the process.

A more detailed description of the process is found in Appendix C.

5.1.1.6 SUSAN Technology

A large European project named SUstainable and SAfe re-use of municipal sewage sludge for Nutrient recovery (SUSAN) is currently in progress. This project goal is to develop a sustainable and safe strategy for nutrient recovery from sewage sludge using thermal treatments (Adam, 2005). This project proposes to use incineration to destroy the organic content of the sludge. Then, a thermo-chemical process will be developed to separate the phosphorus and the heavy metals in the ash. In addition to the technological development, the performance of the fertilizer product and the sustainability of the whole process chain will be studied. Because the process is in such an early stage of development, there are no technical or economic details to report.

5.1.2 New Investigations for Building Material Recovery

In Malaysia, Liew et al. (2004a) evaluated the use of dried sewage sludge as raw material in a clay-making process. They concluded that bricks produced by addition of 10 to 40% (by weight) of sewage sludge were capable of meeting the Malaysian relevant technical standards (physical and chemical properties). A leaching test showed that metals concentration from clay-sludge brick is very low (Liew et al., 2004b). However, bricks with more than 30% sludge are brittle and easily broken (Liew et al., 2004a), indicating that the sludge proportion is a key factor in producing high quality bricks.

Others researchers including Anderson and Skerratt (2003), Balgaranova et al. (2003) and Anderson (2002) have experimentally confirmed that sewage sludge has potential for use as an additive in brick manufacture. However, Anderson and Skerratt (2003) pointed out that variability in the consistency of the raw material used in brick manufacturing may have an impact on the quality of the product. Similarly, Coutand et al. (2006) who evaluated sewage sludge ash as mineral admixture in cement based-material, concluded that it does not satisfy the American and European requirements to be directly considered as a mineral admixture.

At the University of Leeds in the UK, Forth (2007) has recently produced a building block made almost entirely of recycled glass, metal slag, incinerator ash, pulverized fuel from power stations, and biosolids. This material is called “Bitublock”. Bitublock is formed by using bitumen to bind the mixture of waste products together before compacting it in a mold to form a solid block. The block is then heated to oxidize the bitumen. This gives the block the structure of concrete.

5.1.3 ARP Technology for Nitrogen Recovery

ARP is an acronym for the Ammonia Recovery Process, a technology developed by the Battelle Memorial Institute in the U.S., currently commercialized by the ThermoEnergy Corporation.

ARP Technology uses a proprietary resin system to extract a commercial-grade fertilizer (ammonium sulfate) from a nitrogen-rich sludge sidestream. The technology was initially developed to treat the nitrogen-rich effluent after sludge processing with the Sludge-To-Oil

Reactor System (STORS). [The STORS is an energy recovery process that will be described latter in this Chapter, Section 5.2.]

In the ARP process, an ion exchange unit concentrates the ammonia in the influent from approximately 1000 ppm to 15,000 ppm (CERF, 2000). The concentrated ammonia stream is then vaporized. Finally the ammonia gas gets in contact with sulfuric acid and crystallized as ammonium sulfate.

The first ARP pilot plant was constructed at Oakwood Beach Water Pollution Control Plant (WPCP) Staten Island, New York. The pilot plant was tested from September through December of 1998. The evaluation was conducted using centrate produced during dewatering operations of the plant's anaerobically digested sludge. No recent information on the status of the project was identified.

A more detailed description of the process is found in Appendix C.

5.1.4 BIOSOL Process for Low Metal Containing Compost Production

The BIOSOL process was developed by the University of Toronto, Canada. This process is used to remove metals and destroy pathogens from wastewater sludge (e.g. primary sludge).

The process can serve as pre-treatment prior to a composting phase. With BIOSOL, composts are produced with low concentration of metals. An onsite demonstration of this process was conducted at the Greater Moncton Sewerage Commission's (GMSC) facility (LeBlanc et al., 2004).

5.1.5 Volatile Acids Production

Various volatile acids such as formic acid, acetic acid and propionic acid can be produced from sludge. This can be achieved by microbiological (anaerobic) processes, or thermal processes such as wet air oxidation (Rulkens, 2004). It as been shown by Shanableh and Jomaa (2001) has shown that an optimum quantity of volatile fatty acids (VFAs) can be produced at moderate hydrothermal treatment, with temperatures below 200°C. Recovered volatile acids might serves as a chemical feedstock, but no developments in this area were identified.

5.1.6 Bio-Pesticides

Research activities are currently on-going to recover bio-pesticides from sludge. For instance active components (crystal proteins, spores and other factors of virulence) of *Bacillus Thuringiensis* based bio-pesticides were recovered, from secondary sludge, by ultrafiltration by a Canadian research team and U.S. EPA (Adjalle et al., 2007). The secondary sludge originated from the wastewater treatment plant of Communauté Urbaine de Québec, Canada.

5.2 Emerging Technologies for Energy Recovery

5.2.1 Emerging Sludge-to-Biogas Processes

5.2.1.1 Anaerobic Digestion

The Bioterminator$^{24/85}$

Bioterminator$^{24/85}$ is a modification of mesophilic (35°C) anaerobic digestion technology developed by Total Solids Solutions, LLC from research conducted at the University of Louisiana, U.S. The main difference between this technology and the conventional anaerobic digestion process is the mixing pattern. Bioterminator$^{24/85}$ works with laminar plug flow hydraulics as opposed to complete mixing (TA, 2004). This system works on secondary or mixed primary/secondary sludge with a solids concentration of 0.5-5.0% TS.

During the digestion process with BioTerminator $^{24/85}$, a supplemental buffering agent may be fed to maintain the pH. This depends on the characteristics of the raw sludge. Additionally, a low dose of sucrose is fed to the digester as a catalyst.

This technology is reportedly capable of destroying 85% of TS in 24 hours. The detention time of the reactor is 24 hours or less. A maximum of 25% of the digester gas is required for heating to maintain the temperature of the digester to 35°C.

The first full-size installation is scheduled to be constructed in Daphne, Alabama in 2007 (U.S. EPA, 2006).

The short retention time is a significant improvement in anaerobic digestion technology, as it would greatly reduce the size of the anaerobic digestion tank, with ensuing benefits from a land footprint and from a life cycle assessment of reduced demand for construction materials.

A more detailed description of the process is found in Appendix C.

Columbus Advanced Biosolids Flow-through Thermophilic Treatment (CBFT3)

The CBFT3 is a modification of thermophilic anaerobic digestion using a plug-flow reactor. It was developed by Columbus (GA) Water Works and Brown and Caldwell Consultants. While a main objective of the process is to improve pathogen reduction in biosolids, the process incorporates advanced reciprocating engines to produce electricity that supplies 40-50% of the plant electricity needs. The overall energy efficiency of the process is 68-83%. The process is discussed as a Case Study in Section 6.0.

5.2.1.2 Use of Biosolids Pellets for Bio-hydrogen Gas Production

A key challenge to implementing fermentative hydrogen production is the presence of hydrogen-consuming microorganisms in the inoculum source. In laboratory studies, researchers use heat shock treatments, pasteurization or boiling, to kill many of these organisms that are present in natural inoculums.

Heat treatment is only a short term solution, however, if the wastes to be treated include hydrogen-consuming microorganisms. In that case, hydrogen-consuming organisms will re-colonize the system. At full scale, re-colonization would be continuous and ongoing heat treatment of the waste would be impractical and economically infeasible. Heat treatment of an inoculum source that is fed to a hydrogen-producing reactor under a bio-augmentation strategy may be feasible, however, if the inoculum source would be heat treated in any case.

Kalogo and Bagley (2007) demonstrated experimentally that biosolids pellets from anaerobically digested municipal wastewater sludge by drying to greater than 90% TS at 110-115°C for at least 75 minutes could be a practical inoculum source for hydrogen reactors. The experiment has only been conducted at bench scale. The study requires further investigations at pilot scale since the biosolids pellets could serve as an inexpensive seed for ongoing bio-augmentation.

5.2.1.3 Ozonation

Ozonation is an oxidative pre-treatment process that is used to enhance sludge hydrolysis and improve solids biodegradability. During a pilot study in Japan, a mixture of primary and secondary sludges was pre-treated by ozonation and then treated in an anaerobic digester (Goel et al., 2004).

Part of the digested sludge was withdrawn from the digester, ozonated and returned back to the reactor. The sludge retention time of this system was very long (275 days) because of the recycle stream. A long retention time is one the key points of the process for a high removal efficiency (Goel, 2007 - personal communication). This new process scheme resulted in organic solid degradation of 81% while the TS degradation efficiency was 61%. The process produced 36% more energy than a digester treating the same quantity and same sludge without ozonation.

The production of ozone requires a high energy input, however. The required energy during this experiment was higher than the surplus generated, so the negative energy balance is a significant hurdle to overcome.

A more detailed description of the process is found in Appendix C.

5.2.1.4 Pulsed Electric Fields

This technique uses high intensity pulsed electric fields for cell destruction. Kopplow et al., (2004) investigated pulsed electric fields as a pretreatment method prior anaerobic digestion. The experiment showed positive effect on biogas production. It has however a high energy demand compared to other sludge disintegration methods. Neither a pilot scale nor a full-scale demonstration has been conducted yet with the technique.

Banaszak et al (2007) demonstrated at lab and pilot scale that pulsed electrical field pretreatment of waste activated sludge can increase the methane production in anaerobic digester and reduce significantly the amount of sludge for disposal. The increase in methane production is due to the increase in soluble COD. The authors predicted that the net sludge volume for disposal could be reduced by 80%. A commercial full-scale installation of this process has been recently initiated at the Mesa, AZ Northwest Water Reclamation Plant.

5.2.1.5 Enzymatic Hydrolysis

Enzymatic hydrolysis pre-treatment is a method that has also been investigated to enhance energy production from anaerobic digesters. The method involves creating an environment that enhances enzyme activity of the anaerobic bacteria in several reactors in series (Le et al., 2006). The initial enzymatic process tanks operate at 42^{o}C and short detention time to promote acidogenic bacterial growth, while the subsequent process tanks operate at 55^{o}C to promote growth of methanogens (Le et al., 2006). Enzymatic hydrolysis pre-treatment can increase VS destruction by around 10% with concomitant improvements of 24% in biogas production (Werker et al. 2007). Another advantage of the enzymatic process is the production of volatile fatty acids which can be used for biological nutrient removal in wastewater treatment plants (Le and Harrison; 2006). The third advantage of the process is it capability to achieve sludge disinfection at mesophilic temperature (Le, 2007). Full-scale experience undertaken at the Blackburn wastewater treatment plant (UK) has shown that factors such as hydrolysis kinetic, inhibition and probably many other unknown factors can limit the potential for volatile fatty acid production during enzymatic hydrolysis (Werker et al., 2007).

5.2.1.6 Microwave Irradiation

Microwave irradiation is another innovative method that can be used increase waste activated sludge VS solubilization. This method consists of pre-treating WAS at temperatures higher than the boiling point, between 110^{o}C and 175^{o}C (Toreci et al., 2007). The effluent is then treated by a mesophilic anaerobic digester in which up to 30% more biogas can be produced.

5.2.1.7 AFC^{sm} Process

The AFC^{sm} process is described by the manufacturer (PMC BioTec Company, PA, U.S.) as a catalyzed membrane bioreactor with negligible organic sludge production (PMC BioTec, 2008a). The process is able to operate with either aerobic or anaerobic process sludges. As an aerobic membrane reactor, sludge oxidation is accelerated by side-stream hydrolysis of cells. Because it is operated in the 40-75 ^{o}C temperature range, the kinetics of sludge degradation are much higher than typically encountered in normal aerobic digestion. When operated as an anaerobic membrane process (termed the AFC BioFuels^{sm} process), feed sludge are reportedly converted to methane and carbon dioxide and water at an efficiency of 85%, ,also resulting in a reduction in residual sludge disposal costs of up to 75% (PMC BioTec, 2008a).

The technology appears to be emerging from the development stage, although it is difficult to determine the number of operational full-scale facilities. A pilot facility was installed in Coleraine, Northern Ireland for seven months, with reported COD reduction of 97% in the WAS, and total suspended solids destruction of 84% (PMC BioTec, 2008b). The estimated capital cost for a 20 MGD wastewater facility producing 18 dry tons/d of thickened WAS is approximately $US 2.2 million. Operating costs, excluding capital amortization are reported to be on the order of $US100-300per dry ton of organics (PMC BioTec, 2008a).

The literature reports a full-scale industrial facility operating in the aerobic mode on the U.S. East Coast costing $US 33.8 million, but saving $US2.5 million per year in sludge hauling costs (PMC BioTec, 2008a).

5.2.2 Emerging Sludge-to-Oil Processes

5.2.2.1 Sludge-To-Oil Reactor System

Sludge-To-Oil Reactor System (STORS) technology was developed by Battelle Memorial Institute in the U.S. The technology is currently commercialized by ThermoEnergy (ThermoEnergy, 2007). Preliminary research was conducted around 1986 (Molton et al., 1986). In the process, dewatered sludge (about 20% dry solids) is mixed sodium carbonate (Na_2CO_3) prior to charging the mixture to the reactor. At high pressure and elevated temperature, after 1 to 3 hours, the sludge is converted into a fuel consisting of an oil with 90% of the heating value of diesel, and a solid "char" similar to coal. The recovered oil can be used to produce electricity and/or heat using an engine.

There is currently no full-scale installation in operation. A pilot scale unit (20,000 m^3/d) was operated at the Colton MWTP in California. During this project, the reactor was combined with the ARP (previously described above) to recover both biofuel and ammonia. The Colton project began in September 1998 and was concluded in November 2000. The project cost, in 1998 US$, was estimated at 3 million and was funded by the U.S. EPA. Information such as energy input, energy generated, O&M costs for the Colton project is not available in the literature.

The process is complex, requiring many process vessels, resulting in a substantial land footprint. Commercial success may be tied to inclusion of an ammonia recovery process (ARP). Because of operating at high pressure and temperature, energy inputs are anticipated to be significant, resulting in a substantial carbon footprint.

A more detailed description of the process is found in Appendix C.

5.2.3 Emerging Sludge-to-Liquid Processes

5.2.3.1 Super Critical Water Oxidation

Aqua Citrox® Technology

The Aqua Citrox® process was developed in 1995 by Chematur Engineering in Japan. As a SCWO process, the reactor operates at high temperature and pressure. After pre-treating the sludge (up to 15% dry solids) to remove large particles, the feed passes through a high pressure pump. It is then pre-heated using reactor effluent before it enters in the SCWO reactor. The detention time of the reactor is about 1 minute. At the high process temperature, nitrogen contained in the sludge is completely transformed to nitrogen gas.

A pilot-scale unit with a capacity of about 1100 kg per hour was built in around 2000 in Japan. This experiment showed that the removal efficiency of COD and nitrogen by the Aqua Citrox® process was higher than 99.99%. No full-scale commercial installations were identified.

Operation of the Aqua Citrox® requires electricity, natural gas and oxygen (Gidner and Stenmark, 2001). SCWO is a complex process involving high pressures and temperatures, however, there is little chemical use other than the pure oxygen required. The land and carbon footprints are expected to be substantial.

Aqua Reci™ Technology

The Aqua Reci™ Technology is also a SCWO process. It is discussed later in this section because it combines both energy and phosphorus recovery.

A more detailed description of the process is found in Appendix C.

5.3 Processes for Energy and Resource Recovery

Discussion above has reported on either energy recovery or resource recovery, but not an integration of both concepts. In this section, emerging processes that combine energy and resource recovery are discussed.

5.3.1 KTH Two-Stage Acid-Base Leaching Concept

A two-stage process (Hultman, 1999) using both acid and base extractions of phosphorus has been developed by Kungl Tekniska Högskolan (KTH), the Swedish Royal Institute of Technology (Levlin and Hultman, 2004). Sludge from a BNR system is treated in an anaerobic digester to produce biogas (energy) and release phosphorus and ammonium (resources). The digested sludge solids are separated from the liquid phase (supernatant). The latter is used to recover struvite or to produce phosphoric acid by extraction. The digested sludge can be conditioned with heat, pressure, acids, etc. and dewatered. The filtrate from dewatering is further treated to selectively remove toxic metals. Additional energy is recovered from the sludge residue by incineration.

For the phosphorus recovery stage, a low dose of acid (hydrochloric acid, HCl) first dissolves calcium, magnesium and part of the phosphorus contained in the ash. A low dose is necessary to avoid leaching of aluminum (Al^{3+}). Although a high dose of HCl would generate more phosphorus, aluminium would have to be separated from the leachate prior recovering the phosphorus, resulting in greater effort and expense. The leachate from the first step is then further treated with sulfuric acid (H_2SO_4) to produce phosphoric acid (H_3PO_4). Calcium phosphate can then be produced through addition of lime. In the final stage of this system, caustic soda (NaOH) is used to leach the remaining phosphorus in the solids from the first stage.

Although the process has been tested at bench scale (Levlin, 2006; Levlin, 2007), no full-scale application has been reported to date.

The anticipated advantage of the process is a low heavy metal contamination of the final product. A similar concept named SEPHOS, under development in Germany, was described earlier in the report. As with many of the emerging phosphorus recovery technologies, a number of process steps are required to achieve the final recovered product. The process does not require high pressure and temperature, but incineration of the sludge is needed to use the ash for P recovery. The need for incineration and the staging of the processes leads to a significant land requirement.

A more detailed description of the process is found in Appendix C.

5.3.2 Aqua-Reci^{TM} Technology

The Aqua Reci^{TM} Technology was developed in Sweden jointly by Chematur Engineering AB and Feralco AB (Chematur, 2007). The process recovers both energy and a resource material (phosphorus). This technology combines the SCWO process for energy recovery (as described above) used to produce ash, with subsequent chemical leaching of phosphorus from the ash.

Phosphorus is extracted by leaching the ash with hydrochloric acid (HCl), sulfuric acid (H_2SO_4) or caustic soda (NaOH) at a moderate temperature of approximately 90^{o}C . Leaching the ash with a weak NaOH solution can extract 65-90% of the phosphorus from the ash, while 100% of the phosphorus can be extracted with HCl or H_2SO_4 (Stendahl and Jäfverström, 2004). Although acid leaching is more efficient, alkaline leaching gives lower metal contamination in the leachate (Stark, 2005; Levlin et al., 2004).

Pilot-scale experiments of the process were conducted at Karlskoga, Sweden. An energy balance on the the Aqua Reci^{TM} Technology (Stendahl and Jäfverström, 2003) indicated that it should be a significant excess energy producer, over and above the demands for oxygen production and heat and pressure requirements for the SCWO reactor.

A pilot scale Aqua Reci^{TM} unit was tested to recover phosphorus in Karlskoga, Sweden. The test was conducted with digested sludge (15% dry solids) from sewage treatment plants in Stockholm and from Karlskoga sewage treatment plant.

A more detailed description of the process is found in Appendix C.

6.0 International Case Studies

This chapter presents case studies of examples of energy or resource recovery in a consistent manner that allows comparison of the different technologies. The case studies were developed for the report by members of the Global Water Research Coalition.

Case Study #1 – Energy Recovery

Columbus Advanced Biosolids Flow-Through Thermophilic Treatment Columbus Water Works, Georgia, USA

Objective:

This project was first to develop an innovative plug-flow digester used in conjunction with thermophilic anaerobic reactors for improved pathogen destruction. The new process also takes credit for pathogen destruction in continuous feed digesters and shortens the required batch time considerably. An enhancement has been included to optimize the produced biogas to generate heat and power.

Status

This project has been underway since 2002. The project included laboratory-scale studies (now complete) and the construction of a full scale plug-flow prototype completed in 2004. The project is being expanded at the Columbus digestion facility by installing advanced reciprocating engine systems (ARES) to generate power as well as to recover heat from the thermophilic phase.

Background:

Columbus Water Works (CWW) serves Columbus, Georgia, a city located on the Chattahoochee River on the western border with Alabama. Columbus is the third largest city in Georgia and the fourth largest metropolitan area in the state. CWW provides wastewater collection and treatment to 62,000 locations serving a population of approximately 200,000 and recently assumed ownership and operation of the US Army's water and wastewater service at Fort Benning. The project is at the South Columbus Water Resource Facility which currently treats 132,500 m^3/d of sewage.

Description

The Columbus Biosolids Flow-through Thermophilic Treatment (CBFT3) is a conventional, complete-mix, continuous feed thermophilic anaerobic digester followed by a long, narrow plug-flow reactor. Mesophilic digestion is the final stage is used to minimize odors in the wastewater residuals. The process has provisions for using all of the gas from the digesters to generate power to partially offset the facility's electrical consumption. In addition, heat is recovered from the thermophilic biosolids cooling to mesophilic temperatures to heat the raw solids. In a high grade heating loop, heat is supplied by the engine jacket cooling water and exhaust.

In a final heat recovery step, grease trap waste is introduced into the digester feed stream and heated by ARES intercooler and lube oil cooling heat to raise the temperature to thermophilic range. Power generation from biogas is accomplished by the ARES. ARES units were selected from several options which were evaluated based on their low capital cost, high electrical efficiency, and ready availability of high grade heat for heating the digestion processes to thermophilic temperatures. ARES is an upgraded conventional engine technology offered in the US by three vendors, developed with the support of the U.S. Department of Energy and the U.S. National Laboratories.

Digester gas contains contaminants that affect cogeneration equipment. Treatment of the biogas to remove water, siloxanes, hydrogen sulfide and particulates is required. CWW uses a multi-stage gas treatment system consisting of a series of processes to cleanse the gas of these contaminants. The system includes an iron sponge to neutralize hydrogen sulfide, refrigerative drying to remove water, activated carbon unit for siloxane removal, and a fine-media filter for removal of particulates.

Columbus Advanced Biosolids Flow-Through Thermophilic Treatment

Results

- Average volatile solids reduction of 58% (mesophilic digestion only. Thermophilic predicted to be 68 – 70%. Construction of full-scale CBFT3 improvements and power co-generation underway November 2007 – 2008.) The following results are predicted:
    - Average gas production is about 9,200 m^3/d of biogas
    - ARES electrical efficiency of 39%
    - Thermal efficiency of 30% in summer and 42% in winter
    - Combined overall efficiency between 68 – 83%
    - Co-generation expected to produce 1.2 – 1.4 MW electrical power
    - ARES engines can produce 40-50% WWTP power requirements.

Advantages

- Technologies for pathogen destruction and co-generation of heat/electric energy are complementary processes
- Digester gas burns better than landfill gas and lower NOx emissions than natural gas in ARES engines
- Recovered heat fully heats solids process 99% of time

Disadvantages/Limitations

- Contaminants in biogas must be addressed before use
- Multistage gas treatment systems cost more to construct than single stage systems but require lower maintenance

Cost/Pay Back Period

Base payback estimated 9.6 years (no carbon credits and electric energy savings at US$0.055/kWH)

Energy Benefits/Carbon footprint

11,700 metric tons per year of CO_2 reductions calculated as 12.3 gigawatt-hours per year at 2.1 lbs. of CO_2/kWh.

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Vendor Contact

Developed by CWW with:
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Patent rights for the CBFT3 process were given to WERF in 2005.

References

Willis, John, Cliff Arnett, Steve Davis, Jim Schettler, Anup Shah, and Randy Shaw. 2007. “Maximizing Methane – A Georgia treatment plant intends to test a new method for optimizing the use of digester gas to generate power”. *Water Environment & Technology*. www.wef.org/magazine. February issue.

Case Study #2 – Energy Recovery

Energy Interaction with Waste Incineration Plant, Nijmegen, NL

Objective:

The cost of a wastewater treatment plant (WWTP) expansion for full nitrogen and phosphorus removal is enormously reduced by innovative heat exchange with an incineration plant.

Status

Full scale application since November 2000.

Background:

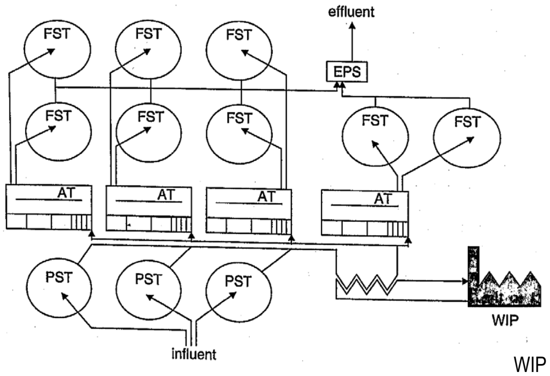
The Nijmegen WWTP in The Netherlands was built in 1982 only for BOD removal and partial nitrification. In the 1990s, several adaptations were carried out to meet new regulations for phosphorus removal. In 2000 a design of an extension and retrofitting of the aeration tanks and secondary clarifiers was made (455,000 pe; 4,530 m^3/h) with effluent requirements of 10 mg/l N and 1 mg/l P.

At the waste incineration plant Nijmegen (WIN), located directly next to the WWTP, heat is used for the production of steam. Electricity is made by expansion of the steam in two turbines. The heat from steam condensation is removed by two air-cooled condensers. On summer days, there was not always enough cooling capacity, with the consequence that the desired amount of waste could not be incinerated.

A win-win situation was created by reducing the incinerator's investment in new air-cooled condensers, increasing their waste processing capacity and producing electricity while reducing the capital and operating cost for the expansion of the WWTP for the required nitrogen removal. Warming part of the primary sedimentation overflow leads to a constant process temperature of 20-25°C. The higher temperature of the influent allowed for smaller tank volume during the expansion for nutrient removal.

Description

A part of the primary sedimentation overflow (25 % of DWF) is heated up and subsequently fed to the aeration tanks. The heat exchange circuit consists of a pump station for pretreated wastewater on the WWTP site, a piping system and a water-cooled condenser on the site of the incineration plant. The required amount of pretreated wastewater is about 1,200 m^3/h, which is always available. The WWTP uses the exchanged heat for warming up the aeration tanks. There is enough heat to create process temperatures higher than 16°C throughout the year and the process temperature can be increased to a maximum 28°C in the summer.



Operation in 2003 WWTP with waste incineration plant (WIP)
(PST=primary sedimentation tank; AT= aeration tank; FST= final sedimentation; EPS= effluent pumping station)

Energy Interaction with Waste Incineration Plant

Results

Figure 2 shows the process temperature in the aeration tank before and after process heating. The COD removal improved (figure 1) and the sludge volume index throughout the year is very stable (80-120 ml/g).

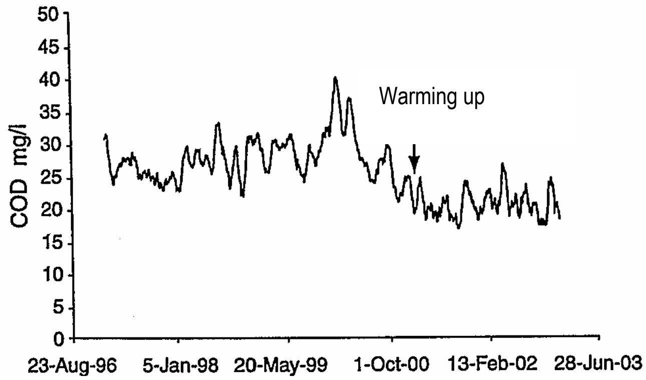


Fig 1: COD effluent concentration form March 1997 until December 2002

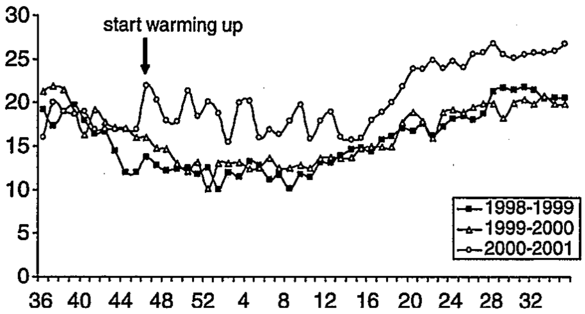


Fig. 2: Process temperature of the aeration tank Measured during three consecutive years

Advantages

- A higher process temperature range (approximately 5 – 6°C higher) throughout the year in the aeration tanks leads to advanced COD removal and P removal.
- The incineration plant has a profit of about US$1,409,141/year due to a greater incineration capacity and its own production of more electricity.

Disadvantages/Limitations

- Besides some clogging problems in the heat exchange circuit in the beginning the system functions very well.
- Limited WWTP's are located near incineration plants or industries with heat surplus.
- A small increase of odor emission is due to a higher process temperature. This odor emission has not caused more odor complaints.

Cost/Pay Back Period

The investment costs for the expansion of the WWTP could be reduced with about US$4,227,573 (30-40%). The reduction of the calculated annual costs is US$352,320/year.

Energy Benefits/Carbon footprint

Not available

Facility Contact

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-
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References

Kruit, J., et.al. 2005. "Innovative energy interaction with a waste incineration plant leads to advanced biological P and COD removal and cost reduction" *Water Science and Technology* 52 (4)

Case Study #3 – Energy Recovery

Cambi Thermal Hydrolysis Pre-Treatment for Mesophilic Anaerobic Digestion, Bran Sands Regional Sludge Centre, UK

Objective:

This project was designed to convert a regional sludge centre from raw sludge drying to advanced digestion and dewatering (with option to dry). The new plant will also have 5 MWs of cogeneration - combined heat and power.

It was also designed to move away from transport of sludge from satellite plants as liquid by road and sea and to transport sludge as raw cake in order to reduce the transport costs and impact of the sludge centre.

Status

The project was developed in 2005 as a case study and rapidly moved to design and construction during 2006/7. The plant is now under design and will be operational by 2009. Running parallel to the study were laboratory trials on thermal hydrolysis, anaerobic digestion and dewaterability that confirmed the vendor’s claimed advantages for this sludge.

Background:

Northumbrian Water Ltd (NWL) treats the sewage for about 4 million people in the north east of England. The project will treat the sludge from 2 million people. The Bran Sands site near Middlesborough has a WWTP designed to treat mainly industrial effluent. On the same site the existing regional sludge treatment centre (RSTC) was designed to dewater and dry raw sludge from NWL. There are 7 dryers of 5 tonne water evaporation capacity. The majority of the sludge was delivered to the site by ship along the coast or by road tanker. The escalating cost of energy and the green credits available for renewable energy suggested anaerobic digestion was appropriate for NWL which has very little anaerobic digestion.

Description

Initially a study was carried out to review the whole of NWL region and to investigate the conversion of shipping storage tanks in a number of locations to digesters. The study showed that there was an advantage in keeping the RSTC as one major digestion centre for the southern half of NWL and converting existing shipping tanks at Newcastle for the northern half of NWL. The study also showed that there was a major advantage in reducing transport costs for tankers and the remaining ship by installing a number of local dewatering plants and transporting the majority of the sludge as cake to the RSTC.

The report also recommended a number of advanced digestion technologies that would maximize the green credit payback and minimize digested solids. For the RSTC the recommendation was for Cambi Thermal Hydrolysis (THP) as a pre-treatment.

A contract for building the green field cake reception and digestion centre was written with a preference for thermal hydrolysis pre-treatment of sludge and an outline design for the facility (alternates were allowed). The contract was awarded in April 2007 and design has begun for the THP and associated works.

Running parallel, the sludge from RSTC was dewatered and thermally hydrolysed in a pilot plant and digested in lab digesters during late 2006 and early 2007. The outputs were to confirm the vendor’s claims for digestion loading, digestion rate, dewaterability and pathogen control amongst others.

Cambi Thermal Hydrolysis Pre-Treatment for Mesophilic Anaerobic Digestion

Results

The key deliverables for the project were: Minimum digestion volume – 5kgVS/m^3/day load rate/15-20 days HRT; Volatiles conversion 60%, Digested sludge dewatering >30% DS and a pathogen free product. The lab trials demonstrated the loading rate, the dewatering at 32% DS and the pathogen free product. The VSR method was not successful (incorrect measurement of hydrolysed sludge DS%) but relative biogas production indicated 60% COD conversion at 20 days.

Advantages

- Feedstock is raw cake – 16% DS, digester fed with 10% DS hydrolysed sludge.
- Continuous automatic process
- Pathogen free product with no regrowth meets tome temp 165°C at 20 minutes
- Biosolids only 60,000 t per year for 40,000raw dry tonnes
- Cake will be low odour, 32% DS (using existing belt presses) Drying not needed
- High conversion of volatiles of high secondary content sludge
- Process and performance demonstrated at other sites since 1995
- High quality biogas for CHP
- Process runs on mainly waste heat

Disadvantages/Limitations

- Needs dedicated operator/maintenance (less than drying though)
- Not quite autothermic on engine waste heat (if all biogas is used in CHP)

Cost/Pay Back Period

6 years

Energy Benefits/Carbon footprint

5 MWs generated. CO2 saving is estimated at 50,000 t CO2, (5,000 avoided electricity, 25,000 avoided natural gas, 20,000 generated electricity)

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References

See Cambi web site for a list of conference papers and reference list www.cambi.com

Case Study #4 – Energy Recovery

Ultrasonic Treatment of SAS prior to Anaerobic Digestion, Cotton Valley, Milton Keynes, UK

Objective:

The aim of this trial was to monitor the effects and expected benefits of disintegrating (floc de-stabilization to full cell lysis) Surplus Activated Sludge (SAS) prior to mesophilic anaerobic digestion. Some of the anticipated benefits included an increase in biogas production and an increase in solids destruction. The trial was held at a full-scale operational plant.

Status

This project began in 2004 and a trial period took place from July to November 2005. The technology is currently being installed at an alternative location in a full-scale installation alongside a CHP generator to utilize the extra biogas.

Background:

Cotton Valley Sewage Treatment Works is in Milton Keynes, Buckinghamshire and treats sludge from site and works in the surrounding area. Cotton Valley treats 9600 tds/yr and a population equivalent of 261,000.

Description

Anglian Water relies heavily on anaerobic digestion to treat sewage sludge and has been researching ways to improve performance for a number of years. This has specifically looked at increasing biogas, to provide energy, using combined heat and power (CHP) systems, and reducing solids to lower disposal costs.

Cotton Valley has two anaerobic digesters and each digester received an average sludge feed of 250m^3 a day, of which 40% consisted of SAS. The digesters operated at an average hydraulic retention time of 14-15 days and were mixed for 18 hours a day using unconfined gas mixing. The ultrasound unit was installed on the secondary sludge feed line to one of the two digesters. The other digester received an equal volume of SAS, which was not treated by ultrasound.

The ultrasonic equipment was a Sonix ultrasound containerized plant. The works sludge make required a 5 stack installation capable of treating up to 150m^3/d of thickened activated sludge. The horns were of radial design and each rated at 3kW continuous duty. The residence time within the “active zone” of the reactor was 2 seconds. The Sonix ultrasound system is used to disintegrate surplus activated sludge that is difficult to digest due to its cellular content.

Ultrasonic Treatment of SAS prior to Anaerobic Digestion

Results

- A 22.4% increase in biogas during the trial period.
- During some periods the increase in biogas was much higher; around 48% in July 2005.
- The impact of ultrasound on solids destruction (DS and VS) was less clear with the test digester only improving percentage destruction for some of the trial period and overall showing little difference in mean DS and VS destruction for the whole period.
- The increase in gas should have resulted from an increase in solids destruction and this may have not been seen due to the sampling occurring too infrequently.
- The parameters that were measured to indicate stability, including alkalinity, volatile fatty acids and ammonia, did not differ significantly between the test and control.

Advantages

- A significant increase in biogas over standard anaerobic digestion.
- An increase in biogas should result from a reduction in VS and therefore total solids.
- Easily retrofitted into an operational sewage works.

Disadvantages/Limitations

- High capital cost
- Upstream process problems can have a knock-on effect on the run-time of the unit.

Cost/Pay Back Period

Payback estimated at 4-7 years

Energy Benefits/Carbon footprint

Based on findings from the trial the Sonix would have a ratio of 1:5.5 for the electricity required to power the Sonix to that from the excess biogas.

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References

Hogan, F. Mormede, S. Clark, P. and Crane, M. 2004. “Ultrasonic sludge treatment for enhanced anaerobic digestion”. *Water Science and Technology,* 50 (9), 25-32.

Case Study #5 – Resource Recovery

GlassPack® Vitrification Technology, USA

Objective:

This technology was developed to vitrify (melt) wastewater solids as a disposal alternative. The technology uses the organic fraction of biosolids as a renewable fuel source to produce an inert glass aggregate product from the inorganic (ash) fraction. The aggregate has multiple beneficial reuse outlets.

Status

This technology is in full-scale operation at two locations:

- Since 1998, Fox Valley Glass Aggregate plant treats about 1,179 MT of wet paper mill sludge per day and produces 66,065 MT of glass per year (Minergy, 2007).
- North Shore Sanitary District (NSSD) Zion, IL

Background:

Minergy Corp. has developed, patented and commercialized vitrification (melting) technologies for the recycling of high volume wastes including municipal biosolids. GlassPack is Minergy's third-generation vitrification process developed to vitrify materials that include biosolids in a process integrated into the wastewater treatment plant. GlassPack is a patented closed-loop oxygen-based vitrification process. Pre-dried biosolids (<15% moisture) are melted at temperatures between 1,330 and 1,500^{o}C. At these temperatures the ash fraction forms a molten glass that is quench-cooled to form glass aggregate.

Description

The North Shore Sanitary District (NSSD) uses the GlassPack® vitrification process to treat biosolids from their three wastewater treatment plants. The sludge quantity is approximately 181 MT per day (66,000 MT per year) at 18% dry solids (Minergy, 2007). The NSSD project started operation in September 2006.

The configuration of the GlassPack® Technology at NSSD is depicted in Figure 1. GlassPack® is a closed-loop system which requires pure oxygen to enhance the waste combustion. At NSSD, oxygen with 90% purity is used. The wet sludge, with 17 to 20% solids, is dried in a fluidized bed dryer before being injected along with oxygen, into a melting chamber. The organic content is combusted, releasing heat energy at temperatures of between 1200^{o}C and 1400^{o}C. The ash component melts at the high temperatures into molten glass. The separation of molten glass and exhaust gas occurs by gravity draining of the glass into a quench tank. The gas is exhausted exterior to the melting unit. The exit gas temperature depends upon the heat recovery technology used.

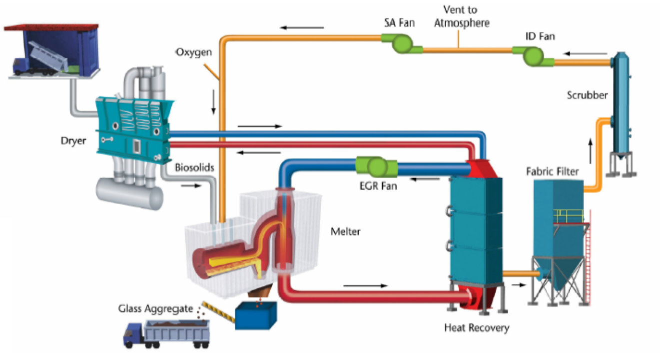

Figure 1. Configuration of the NSSD GlassPack® closed-loop system (**From Minergy, 2007**)

North Shore GlassPack® Technology Performance

Results

- About 17,400 MJ per hour of heat is required to operate the fluidized bed dryer.
- The dryer removes about 5.81 MT of water per hour and produces 32 MT of granulate per day at 92% dry solids content.
- The melter thermal input capacity is 27,432 MJ /hr and produces about 19,900 MJ/hr of exhaust gas. Heat energy is recovered from the flue gas using a thermal oil heat transfer system.
- Approximately 17,400 MJ/hr of heat is transferred by the thermal oil system to the dryer to pre-dry the biosolids. Under normal operating conditions, the dryer does not require additional fuel to complete the drying of the biosolids.
- The plant does not produce electricity.
- The plant produces up to 6.8 MT of glass aggregate per day. The glass aggregate produced at NSSD is approved for beneficial reuse by the Illinois EPA and Wisconsin Department of Natural Resources. Local municipalities use the aggregate as fill in utility trenches.

Advantages

- Can eliminate need to co-fire fuel to achieve vitrification.
- Provides significantly reduced air emissions when compared to other incineration technologies.
- Eliminates ash disposal as the glass aggregate has many reuse options.
- Modularized units reduce construction costs and field installation schedules.

Disadvantages/Limitations

- Natural gas or other external fuel source is needed during start-up.
- Requires a source of >90% pure oxygen on-site.

Cost/Pay Back Period

Capital cost of the plant is about US$32 million and the O&M costs are approximately US$126 per wet MT. Based on the capacity of the plant, the capital cost investment is approximately US$484 per wet MT of annual treatment capacity.

Energy Benefits/Carbon footprint

Offsetting the consumption of natural gas to complete drying, GlassPack saves more than 3.5 million m^3 or more than 140 GJ/year. Use of heat for drying saves in excess of 6,600 MT/year of carbon emissions.

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References

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www.minergy.com

Case Study #6 – Resource Recovery

Phosphorus Recycling from Iron-poor Wastewater Sludge, NL

Objective:

Explore the market potential and technical feasibility of phosphorus recovery from ash of incinerated sludge from wastewater treatment plants.

Status

Feasibility study and full scale pilots.

Background:

Sewage sludge from wastewater treatment plants (WWTPs) is incinerated in the Netherlands in centralized sludge incineration installations. The phosphorus present in the sludge is not re-used and is thus permanently removed from the natural cycle. Natural stocks of phosphorus rock are limited and will eventually be insufficient to meet demand. Closing the phosphorus cycle is becoming more and more important.

The goal of phosphorus producer, Thermphos in the Netherlands, is eventually to replace about 20 percent of the raw mineral material by recycled phosphorus. Phosphorus recovery from WWTP incineration ash could qualify for this purpose. In the past a great deal of research has been done on recovery of phosphorus from the water treatment process in WWTPs, but with mixed results. Of the sludge processed by Slibverwerking Noord-Brabant (SNB), 20-25 % currently meets the criteria for phosphorus recovery (about 25,000 ton/year of dry matter). This sludge has sufficiently low iron content with sufficiently high phosphorus content to enable an ash to be produced with an average Fe/P molar ratio of less than 0.2. An important precondition is that this sludge can be treated separately in the current installation. The future market potential requires an inventory of iron-poor and phosphorus-rich sludge in the Netherlands of at least 50,000 dry tons. Moreover studies must determine the potential at the sludge incineration plants DRSH and SNB. DRSH and SNB are responsible for the processing approximately half of the Dutch sewage sludge and can possibly exchange suitable sludge flows.

Description

Subjects of the research:

1. Large-scale research into the possibilities of employing sludge incineration ash as raw material for phosphorus production.
2. Inventory the market potential of iron-poor and phosphorus-rich sludge in the Netherlands of at least 50,000 tons dry matter. The potential is possibly already present at SNB and DRSH.
3. Identify future trends which could affect the market for iron-poor sludge.
4. Determine consequences of replacing ferriferous additives by aluminous additives and/or waste streams in order to increase the volume of available iron-poor sludge.
5. Investigate phosphorus recovery via precipitates from partial streams of WWTPs.

Results

- Tests conducted in 2006 and 2007 by SNB and Thermphos have shown that it is technically feasible to employ iron-poor sludge incineration ash as raw material for phosphorus production. Recovery in mono-incineration of sewage sludge is promising provided sufficient suitable sludge is available.
- In the Netherlands, phosphorus removal at WWTPs takes place by: dosage of iron or aluminium salts, biological phosphorus removal and combinations of both. The type of sludge processed by SNB and DRSH appears to be sufficiently representative for the Dutch situation. Within DRSH and SNB, the quantity of sludge suitable for processing by Thermphos is calculated to be 41,000 tons dry matter per year. For the whole of the Netherlands, is this approximately 82,000 tons dry matter per year. Potentially, there seems to be sufficient suitable sludge available. For confirmation, better insight into the iron balance in specific local situations is needed. No use of iron salts at the WWTP is no guarantee of suitable Fe/P ratio in the sludge ash, because of the location-specific background concentration of iron.

Phosphorus Recycling from Iron-poor Wastewater Sludge, NL

- It can be concluded that the projected future trends make the proposed application of recovery of iron-poor sludge all the more promising. The application of iron salts for additional phosphorus removal with add-on technology presents uncertainty. In the mid term and long term, no significant changes are expected in the loading, in the production of sewage sludge and in the influent phosphorus load to WWTPs.

  There is a tendency for biological phosphorus removal to be employed more often in order to reduce the consumption of chemicals. In addition, it is recognized that biological phosphorus removal has a favourable influence on the total sludge chain (energy consumption and sludge final processing costs). Due in part to the Water Framework Directive, stricter effluent requirements for phosphorus are expected. This will largely result in additional phosphorus removal with add-on technology such as sand filters.

- Replacing iron-based additives by aluminium-based additives at WWTPs would not, in general, require any modifications of equipment or infrastructure. There is no great disadvantage in this replacement. It is not, however, possible to replace iron by aluminium in all cases. A major advantage of aluminium is the positive effect on the sludge settling in the WWTP. A drawback is that aluminium cannot be used for H_2S bonding. From a cost point of view, commercial aluminium products are more expensive than iron products. On the other hand, there are many aluminium residual products available that are considerably cheaper than commercial iron products which can be universally used for phosphorus fixation. A requirement for phosphorus recovery is that there are sufficient good quality aluminium residual products available locally.
- It is expected that separate collection of phosphorus precipitates from water treatment will not outweigh the yields of phosphorus recovery from incineration ash of the sludge.

This study concluded that there is sufficiently technical and economic basis for recovery of phosphorus from ash of incinerated sludge. It is recommended to:

- publicise the results from the study;
- do detailed investigations into obtaining/exchanging of suitable iron poor sludges (logistical);
- carry out iron analyses of sludge and influent at specific WWTPs to gain more insight in the iron balance;
- examine the availability of aluminium residual products more closely.

Advantages	**Disadvantages/Limitations**
• Reuse of phosphorus	• Limited by current insight in iron balance and Fe/P proportion of the sludge necessary at specific WWTPs • Difficult to influence the background iron concentration in the sewer systems • Possibly logistical improvement sludge processing necessary
Cost/Pay Back Period	**Energy Benefits/Carbon footprint**
• Less than a year	Not available
Facility Contact	**Vendor Contact**
N.V. Slibverwerking Noord-Brabant Middenweg 38; 4782 PM Moerdijk (NL) korving@snb.nl	Not applicable

References

STOWA 2007-23. Fosfaathergebruik uit ijzerarm slib van rioolwaterzuiveringsinrichtingen. ISBN. 978.90.5773.372.7

7.0 Triple Bottom Line Assessment

7.1 Introduction

The Triple Bottom Line (TBL) standard for urban and community accounting has become the dominant approach to public sector full cost accounting. United Nations standards apply to natural capital and human capital measurement as required by TBL, e.g. the ecoBudget standard for reporting the ecological footprint or the human demand on natural resources. This is a measurement of the goal of sustainability. TBL captures an expanded spectrum of values and criteria for measuring organizational (and societal) success: economic, environmental and social.

In this chapter, the TBL approach is used evaluate the technologies reviewed in the previous sections of the report. The goal of this assessment is to identify potential sustainable technologies. A sustainable technology is defined here as one which is economically affordable, socially acceptable and environmentally friendly.

The TBL results presented in this chapter are the consultant team's evaluation, based on the information identified in literature. The assessments are subjective by nature. The limits of the assessment conducted here are also discussed. Relative rankings in the evaluation should not be used in final decision-making because many key pieces information are missing for some of the technologies, and site-specific factors need to be included in a final assessment.

7.2 The TBL Approach

The TBL assessment of alternative technologies is usually based on a set of criteria that is defined for each of the following three elements: economic, social and environmental. A scoring system is then used to measure how each alternative achieve the criteria.

The assessment can be evaluated by assuming that all three elements have an equal weight (base case). It can also be conducted by varying the weight between the TBL elements. This approach requires more detailed information about the alternatives being compared, as well as a way to reflect the values which the population considers each different elements to have. Element weighting is easier to apply to a limited number of alternatives rather than a large number.

Based on the information collected during the literature survey for the technologies described in this work, it appears that there are many knowledge gaps. It is therefore difficult to base the TBL assessment on the conventional approach. The approach used in this subjective assessment includes two parts.

The first part is based on a number of factors that were used as indicators of the social, economic and environmental performance of the technologies. The indicators considered for the assessment are summarized in Table 7-1 for the energy recovery technologies and in Table 7-2 for the resource recovery technologies. The original costs data were converted in 2007 US$ according to the cost index reported by Sahr (2007).

The second part consisted in attributing a value of “+” or “-”, to indicate the performance of a given technology relative to a certain criteria. A positive mark means that the technology achieves well the criteria while a negative mark indicates a poor achievement. A mark of zero was assigned when the nature or value of the indicator was unknown (not found during the literature survey). The selected criteria for each element of the TBL assessment for both energy and resource recovery are summarized in Table 7-3. For each element five main criteria were selected. All criteria were considered equally important, i.e. of equal weighting. The three elements of the TBL were also considered equally important.

The evaluation was based on the nature and/or the perceived value of the indicators. For example, to assign a mark to a certain energy recovery technology, relative to a criteria (e.g. *SG.3.1. Meet minimum land requirement*), Table 7-1 was used to identify the indicators (e.g. *Technology complexity-Number of unit processes*) that can at the best tell if the technology responds to the criteria or not. For evaluation of resource recovery, Table 7-2 was used to identify the indicators.

Table 7-1. Indicators for the assessment of social, economic and environmental performance of the energy recovery technologies.

Process Category	Technology	Technology	Application Full-scale	Feed Concentration	Main Reactor Detention Time	Technology Chemical Use	Operating Temperature	Operating Pressure	Technology Complexity
	Type	Status	*Number*	*% Dry Solids*	*Hours*	*Type*	*°C*	*Bars*	*Unit process*
Sludge-to-Biogas									
Anaerobic digestion	Bioterminator	Emerging	No	0.5-5	24 or less	Sucrose, Buffer	35	*Not Applicable*	1
Thermal hydrolysis	Cambi®	Established	>10	16	0.5	*Not Applicable*	160-180	6	4
Thermal hydrolysis	BioThelys®	Established	2	>10	0.5-1.0	*Not Applicable*	150-180	8	2
Cell destruction	MicroSludge™	Established	2-demo	5-10.0	N.A.	NaOH	*Not Applicable*	1200	7
Cell destruction	Ultrasonic	Established	9	10.0	N.A.	*N.A.*	N.A.	N.A.	2
Cell destruction	Ozonation	Emerging	No	36.60	N.A.	Ozone	N.A.	N.A.	2
Cell destruction	Pulse Electric	Emerging	No	N.A.	N.A.	N.A.	N.A.	N.A.	2
Sludge- to-Syngas									
Gasification	Kopf	Established	1	32[a]	0.5	oxygen or air	900	N.A.	7
Gasification	EBARA	Established	6	N.A.	N.A.	oxygen or air	500-800	N.A.	5
Incineration	Thermylis® HTFB	Established	14	15-35	N.A.	oxygen or air	840	N.A.	5
Sludge- to-Oil									
Pyrolysis	EnerSludge™	Established	1	25[a]	N.A.	No oxygen	450	0.015	4
Pyrolysis	SlurryCarb™	Established	1-demo	10-31.0	N.A.	No oxygen	270-325	N.A.	7
Hydrothermal	STORS	Emerging	No	20	1.0	Na2CO3	215-315	114-148	2
Sludge- to-Liquid									
SCWO	Aqua Reci®	Emerging	1-demo	3	0.017	Pure oxygen	374	220	6
SCWO	Aqua Citrox®	Emerging	No	15	N.A.	Pure oxygen	400-600	250	5
SCWO	Athos®	Established	5	N.A.	N.A.	Pure oxygen	250	50	4

[a] Feed concentration before drying

N.A. indicates information is not available

Table 7-1. (continued) **Indicators for the assessment of social, economic and environmental performance of the energy recovery technologies.**

Process Category	Technology Type	Total Energy Input	Total Energy Output	Capital Cost at data year	Capital Cost at data year	Capital Cost in 2007	O&M Costs at data year	O&M Costs at data year	O&M Costs in 2007	Fate Final Residue
		kWh/dry MT	*kWh/dry MT*	*US$/dry MT.y*	*Year*	*US$/dry MT.y*	*US$/dry MT*	*Year*	*US$/dry MT*	
Sludge-to-Biogas										
Anaerobic digestion	Bioterminator	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	N.A.
Thermal hydrolysis	Cambi®	0.3	5.972	650-1700	2005	677-1775	396	2000	469	Agriculture
Thermal hydrolysis	BioThelys®	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Disposal
Cell destruction	MicroSludge™	502	1358	488	2006	499	75-131	2006	77-134	N.A.
Cell destruction	Ultrasonic	141	N.A.	533-842	2006	545-861	20-25	2006	20.45-25.56	N.A.
Cell destruction	Ozonation	1923	1736	N.A.	N.A.	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	N.A.
Cell destruction	Pulse Electric	N.A.	N.A.	N.A.	N.A.	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	N.A.
Sludge-to-Syngas										
Gasification	Kopf	100	1400	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Asphalt
Gasification	EBARA	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Glass granulate
Incineration	Thermylis® HTFB	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Ash treatment
Sludge-to-Oil										
Pyrolysis	EnerSludge™	120	1966	1640-1954	2007	1640-1954	86-113	2007	86-113	Brick industry
Pyrolysis	SlurryCarb™	712	758	1063-1772	2006	1087-1812	101-104	2006	103-106	fuel in cement
Hydrothermal	STORS	1394-1410	1480-1898	390-4378	2007	390-4378	38-357	2007	38-357	fuel production
Sludge-to-Liquid										
SCWO	Aqua Reci®	738	3167	740	2004	799	230	2004	248	P-recovery
SCWO	Aqua Citrox®	1500	N.A.	158	2001	182	140	2001	161	N.A.
SCWO	Athos®	900-1200	1,680	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Brick industry

N.A. indicates information is not available.

Table 7-2. **Indicators for the assessment of social, economic and environmental performance of the resource recovery technologies.**

Technology	Technology	Application	Feed	Main Reactor	Technology	Operating	Operating	Technology
	Status	Full-scale	Concentration	Detention Time	Chemical Use	Temperature	Pressure	Complexity
Type		*Number*	*% Dry Solids*	*Hours*	*Type*	*°C*	*Bars*	*Unit Process*
P from sludge								
KREPO	Emerging	1-demo	5	N.A.	H2SO4, NaOH, FeCl3, Polymer, Mg(OH)2	100-110	3.6	8
Seaborne	Emerging	1-pilot	N.A.	N.A.	H2SO4, S, NaOH, Mg(OH)2	N.A.		10
Aqua-Reci®	Emerging	1-demo	15	N.A.	HCl, pure oxygen	80-90	220	9
Kemicond™	Emerging	1-demo	2.0-3.0	0.67-1.0	H2SO4, H2O2, Polymer	25	N.A.	5
P from Ash								
BioCon	Emerging	1-pilot	N.A.	N.A.	H2SO4, Ion exchanger	N.A.	N.A.	6
SEPHOS	Emerging	No	N.A.	N.A.	H2SO4, NaOH, Ca2+	N.A.	N.A.	8
P from Side-stream								
Crystalactor®	Established	3	N.A.	N.A.	Ca(OH)2	N.A.	N.A.	1
Phostrip©	Established	4	N.A.	N.A.	Ca(OH)2	N.A.	N.A.	2
Nitrogen Recovery								
ARP	Emerging	1-pilot	N.A.	N.A.	H2SO4, ZnSO4	N.A.	N.A.	7
Building Material								
Thermal solidification-ALWA	Established	>1	30	N.A.	Not Applicable	1050	N.A.	4
Thermal solidification-Slag	Established	>1	30	N.A.	Not Applicable	1500	N.A.	3
Thermal solidification-Brick	Established	>1	30	N.A.	Not Applicable	1000	N.A.	2
Vitrification-GlassPack®	Established	1	17-20	N.A.	Not Applicable	1330-1500	N.A.	5

Table 7-2. (continued) **Indicators for the assessment of social, economic and environmental performance of the resource recovery technologies.**

Technology	Total Energy input	Total Energy output	Capital Cost	Capital Cost	Capital Cost	O&M Costs	O&M Costs	O&M Costs	Final	Product
			at data year	at data year	in 2007	at data year	at data year	in 2007	Product	Usefulness
Type	*kWh/dry MT*	*kWh/dry MT*	*US$/dry MT.y*	*Year*	*US$/dry MT.y*	*US$/dry MT*	*Year*	*US$/dry MT*		
P from sludge										
KREPO	463	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Ferric P	Fertilizer only
Seaborne	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	MAP [c]	Fertilizer only
Aqua-Reci®	738	3167	946	2003	1048	N.A.	*Not Applicable*	*Not Applicable*	Calcium P	P. industry [e]
Kemicond™	187.5	N.A.	198	2007	198	53-98	2007	53-98	Ferric P	Fertilizer only
P from Ash										
BioCon	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	H3PO4	P. industry [e]
SEPHOS	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Calcium P	P. industry [e]
P from Side-stream										
Crystalactor®	N.A.	N.A.	10.95 [b]	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Calcium P	P. industry [e]
Phostrip©	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Calcium P	P. industry [e]
Nitrogen Recovery										
ARP	N.A.	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Ammonia	C. industry [f]
Building Material										
TS [a]-ALWA	1856	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	ALWA [d]	Construction
TS [a]-Slag	1658	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Slag	Ceramic
TS [a]-Brick	2101	N.A.	N.A.	*Not Applicable*	*Not Applicable*	N.A.	*Not Applicable*	*Not Applicable*	Brick	Construction
GlassPack®	5715	3626	4281	2007	4281	650	2007	650	Glass	Road

[a] Thermal solidification, [b] expressed in US$ per kg P removed, [c] Magnesium ammonium phosphorus, [d] Artificial lightweight aggregates, [e] Phosphate industry, [f] Chemical industry

Table 7-3. Selected criteria for each element of the TBL assessment (energy recovery and resources recovery).

SOCIAL	ECONOMIC	ENVIRONMENTAL
SG.1.0: Protect Public Health and Safety	***EcG.1.0: Technology Status***	***EnG.1.0: Greater Energy Efficiency (input-output)***
SC.1.1. Minimize Chemical Use and Storage	*EcC.1.1. Established technology with commercial application*	*EnC.1.1. Technology has lowest net energy input*
SC.1.2. Minimize transport of chemical	*EcC.1.2. At least 5 commercial applications*	*EnC.1.2. No or less use of fossil fuel for upstream raw material production*
SG.2.0: Minimize Community Disruption	***EcG.2.0: Affordability***	***EnG.2.0: Upstream and On-site Emissions***
SC.2.1. No detectable odours	*EcC.2.1. Affordable technology*	*EnC.2.1. Reduce emission of GHG*
SC.2.2. Less visual impact	*EcC2.2. Reduce sludge volume for transportation off-site*	*EnC.2.2. Reduce emission of air pollutants*
	EcC.2.3. Easiness for operating and maintenance	
SG.3.0: Land Requirement		***EnG.3.0: Secondary Waste Production***
SG.3.1. Meet minimum land requirement		*EnC.3.1. Less secondary waste*

7.3 TBL Evaluation

7.3.1 Energy Recovery

This TBL assessment is a subjective evaluation undertaken with limited access to firm data and site-specific circumstances. The rankings should be considered as general indicators, but not serve as a definitive basis for process selection.

The results of the TBL assessment for energy recovery (Table 7-4) suggest that technologies used to enhance biogas production during anaerobic digestion of sewage sludge, including Biothelys, Cambi, and Ultrasonic sludge treatment are the most sustainable. These three technologies have four factors in common: 1) they are established technologies with multiple full-scale commercial applications around the world, 2) no chemicals are used and, 3) operation is relatively less complex than competing processes. They operate with a relatively low temperature compared to the sludge-to-Syngas technologies. Their costs are also competitive when compared with the sludge-to-oil technologies.

The sludge-to-Syngas technologies were evaluated as being somewhat less sustainable than the three Biogas production enhancement processes. Although these sludge-to-Syngas technologies have been commercially demonstrated, they are relatively complex. No information regarding their costs was found during the literature survey.

By this assessment, the least sustainable processes are either under development or in the initial phase of full-scale operation. As a result, the evaluation may not fairly reflect operation if and when the technology is mature. The processes include MicroSludge, STORS, Pulse Electric, Aqua-Reci, Aqua Citrox and Ozonation. In addition, most of the technologies require use of potentially hazardous chemicals as acids or caustics.

7.3.2 Resource Recovery

The results of the TBL evaluation of the resource recovery technologies are shown in Table 7-5.

The assessment indicates that in general, phosphorus recovery from wastewater sludge sidestreams and production of building materials have a relatively higher triple bottom line than do processes for recovery of phosphorus from sludge or ash, or nitrogen recovery as ammonia. Crystalactor® and PhoStrip© use calcium hydroxide, a relatively benign chemical as chemical for phosphorus recovery. Most technologies for phosphorus recovery from sludge or ash use sulfuric acid. The main drawbacks associated with the use of chemicals are health risks and life cycle impacts (upstream energy use and emissions). These drawbacks have been discussed in detail in Section 4.4.3 (Social Feasibility).

Table 7-4. Social, economic and environmental performances of the energy recovery technologies.

	SOCIAL PERFORMANCE					ECONOMIC PERFORMANCE					ENVIRONMENTAL PERFORMANCE				
Technology	***SC.1.1.***	***SC.1.2.***	***SC.2.1.***	***SC.2.2.***	***SC.3.1.***	***EcC.1.1.***	***EcC.1.2.***	***EcC.2.1.***	***EcC.2.2.***	***EcC.2.3.***	***EnC1.1.***	***EnC1.2.***	***EnC.2.1.***	***EnC.2.2.***	***EnC.3.1.***
Type	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***	***Mark***
Sludge- to-Biogas															
Bioterminator	-	-	0	+	+	-	-	+	0	+	+	+	+	+	+
Cambi™	+	+	-	0	0	+	+	-	+	0	+	+	+	+	+
BioThelys®	+	+	+	+	+	+	+	0	+	+	0	+	+	+	+
MicroSludge™	-	-	+	-	-	+	+	+	+	-	+	-	-	-	+
Ultrasonic	0	0	+	+	+	+	+	+	+	+	+	0	0	0	+
Ozonation	0	+	+	+	+	-	-	0	+	+	-	+	+	+	+
Pulse Electric	0	0	0	+	+	-	-	0	0	+	0	0	0	0	+
Sludge-to-Syngas															
Kopf	+	+	+	-	-	+	0	+	+	-	+	+	+	+	+
EBARA	+	+	+	-	-	+	+	+	+	-	-	+	+	+	+
Thermylis® HTFB	+	+	+	-	-	+	+	-	+	-	0	+	+	+	+
Sludge-to-Oil															
EnerSludge™	+	+	+	-	-	+	-	-	+	0	+	+	+	+	+
SlurryCarb™	+	+	+	-	-	+	0	-	+	-	0	+	+	+	+
STORS	-	-	+	+	+	-	-	0	+	+	+	-	-	-	+
Sludge-to-Liquid	+	+	+	-	-	+	-	-	+	0	+	+	+	+	+
Aqua Reci®	+	-	0	-	-	-	-	+	+	-	+	+	+	+	+
Aqua Citrox®	+	-	0	-	-	-	-	+	+	-	0	+	+	+	+
Athos®	+	-	0	-	-	+	+	0	+	0	+	+	+	+	+

Table 7-5. Social, economic and environmental performances of the resources recovery technologies.

Technology	Social Criteria					Economic Criteria					Environmental Criteria				
	SC.1.1.	***SC.1.2.***	***SC.2.1.***	***SC.2.2.***	***SC.3.1.***	***EcC.1.1.***	***EcC.1.2.***	***EcC.2.1.***	***EcC.2.2.***	***EcC.2.3.***	***EnC1.1.***	***EnC1.2.***	***EnC.2.1.***	***EnC.2.2.***	***EnC.3.1.***
Type	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark	mark
P from sludge															
KREPO	-	-	+	-	-	-	-	0	+	-	+	-	-	-	+
Seaborne	-	-	+	-	-	-	-	0	+	-	0	-	-	-	+
Aqua-Reci®	-	-	0	-	-	-	-	+	+	-	+	-	+	+	+
Kemicond™	-	-	+	-	-	-	-	+	+	-	+	-	-	-	+
P from Ash															
BioCon	-	-	+	-	-	-	-	0	+	-	0	-	-	-	+
SEPHOS	-	-	+	-	-	-	-	0	+	-	0	-	-	-	+
P from Side-stream															
Crystalactor®	-	-	+	+	+	+	+	-	+	+	0	-	-	-	+
Phostrip©	-	-	+	+	+	+	+	0	+	+	0	-	-	-	+
Nitrogen Recovery															
ARP	-	-	+	-	-	-	-	0	+	-	0	0	-	-	+
Building Material															
TS [a]-ALWA	+	+	+	-	-	+	+	-	+	-	-	-	-	-	+
TS [a]-Slag	+	+	+	+	+	+	+	-	+	+	-	-	-	-	+
TS [a]-Brick	+	+	+	+	+	+	+	-	+	+	-	-	-	-	+
GlassPack®	+	+	+	-	-	+	+	-	+	-	-	-	-	-	+

[a] Thermal solidification

Overall, the TBL evaluations obtained by the technologies available for resource recovery are low. This reduced performance is mainly due to the use of chemicals, process complexity, land requirements, and the fact that most of the technologies are not yet established. Except for the technologies for building material production, all the other technologies are still in their development phase.

7.4 Limits of the TBL Evaluation

The TBL assessment presented in this work should not be used as final decision to conclude that one technology is systematically better than another one. This is justified by several reasons.

The TBL approach at this level of assessment has a certain level of subjectivity, because of some of indicators used in the evaluation are based on qualitative descriptions of behavior rather than quantitative numbers.

In the present study, many key pieces of information were not readily available for some of the technologies. In some cases, assumptions were made to estimate important indicators such capital and O&M costs, and energy outputs. Use of assumptions involves uncertainties. A typical example found in this study is the assumption that 95% VS could be reduced by the MicroSludge^{TM} process. The consequence of this assumption is a very high production of energy. Another example is that many capital and O&M costs were not available, and thus a neutral value of zero was assigned to the process. With real data, the assigned scores could move up or down relative to other processes for which the data were available

Due to the gaps in the data collected, the total energy inputs and outputs were considered in the TBL assessment. However, some of the technologies produce only heat (e.g. those using super critical oxidation process), while others produce heat and electricity. In wastewater treatment the two products do not have exactly the same value. Although some heat is necessary for process and building maintenance, electricity remains the main energy source required for the operation of the plants.

Additionally, due to the gaps in the data collected, the TBL assessment did not take into account the amount of resources (e.g. phosphorus) that can be recovered by technologies. Another limit of the TBL assessment is that process complexity was established based on the number of unit processes only.

Lastly a number of site-specific considerations that would be part of a formal TBL evaluation could not be addressed. Factors such as federal, state and local government policy, geography, social values, local industry considerations, state of municipal infrastructure, local and wider environmental status, and formal process life cycle analysis, among others, would affect the final evaluations. As noted above, the TBL evaluation is more suited to a limited number of pre-screened processes, rather than as a broadly-encompassing assessment.

7.5 This Report in Wider Context

Although the playing field for this report was defined as energy and nutrient recovery from wastewater solids, this focus is narrow by necessity of time constraints. For true sustainability of wastewater treatment, the field should be expanded using either a cradle-to-grave or cradle-to-cradle approach, which is provided by life cycle or triple bottom line assessments.

In the larger context, issues that could not be included within the boundaries of this report can then be examined. Significant environmental and social issues affecting total wastewater treatment, and indeed the urban water cycle, such as the growing impact of greenhouse gas (GHG) emissions and climate change, or dwindling resources such as mineral phosphate fertilizer, can be evaluated more thoroughly.

With respect to environmental improvements due to wastewater treatment, a number of paradoxes abound. Effluent quality requirements grow ever more stringent, to the extent that reverse osmosis and/or ozone are now being employed; such processes consume substantially more energy to reduce or eliminate contaminants discharged in treated effluents (Monteith et al, 2007). The increased energy use results in greater emissions of GHGs and other criteria pollutants from the electricity generating plants (see Sahely et al., 2005, for example). This is certainly true in developed nations, and will become even more if lesser developed nations adopt existing Western wastewater treatment technologies.

Optimization of specific process operations, if narrowly implemented, may result in overall net damage to the environment. Optimization is typically implemented to comply with effluent regulations, to extend process capacity and to avoid bottlenecks, or to save operating or capital costs. Consider for example a facility that adds a coagulant such as alum to improve removal of suspended solids and phosphorus for effluent quality compliance. The volume of solids for treatment and disposal becomes larger. The dewaterability of the solids is likely to decrease, necessitating aids such as ferric salts and/or polymer. The greater loading of solids for disposal may require additional truckloads for disposal to land application sites or landfills, resulting higher emissions of GHGs and particulates, and greater wear of roads. A procedure is needed to weigh the costs of compliance with water quality with overall life cycle and risk assessments.

Another paradox involves the increasing adoption of biological nutrient removal (BNR), and denitrification in particular, by wastewater facilities in developed nations. With use of anoxic and anaerobic processes, however, the potential for release of nitrous oxide becomes significantly greater (Ahn et al., 2007; Kampschreuer et al., 2007). Nitrous oxide has a global warming potential approximately 300 times greater than that of carbon dioxide (Burrowes et al., 2007). BNR processes also involve greater pumping of recycle streams to the anaerobic and anoxic stages. Pumps and air blowers are the largest energy consumers in wastewater treatment.

With rising energy costs and climate change concerns, recovery of energy at wastewater treatment processes has grown significantly in recent years. In attempting to extract more energy as methane from wastewater solids (principally waste biological solids), chemical and energy use is required. Based on life cycle assessment, the energy required to extract, process, and transport the chemical additives should be included in decisions to adopt a technology, and to determine if such a process is the most environmentally benign way to proceed. Similarly, processes that require energy, either as thermal, sonic or electrical inputs, should be evaluated for net energy and emissions of GHGs or other contaminants. The assessment should determine whether these energy recovery processes are optimized only for energy recovery, or whether they may ameliorate one issue, at the environmental expense elsewhere in the treatment plant, or even outside the treatment plant boundary (i.e., do they have a net positive or negative environmental impact)?

Many processes that are designed to recover materials (nutrients, building materials) from residual wastewater solids also require the use of energy or chemicals. Such processes are subject to the same consideration as the energy recovery processes discussed above.

Following such considerations, there is an apparent need for a paradigm shift (see for example, Leverenz and Tchobanoglous, 2007) in management of the urban water cycle, from a mind-set of disposal to one of reuse and recovery. Points to consider in this paradigm shift can encompass:

- *The need for delivery of potable water for domestic sanitary (toilet) purposes and subsequent transport to centralized wastewater treatment facilities* (*WWTFs)* (considerations include energy costs for treating and pumping potable water, life cycle costs for water treatment and sizing of distribution mains, life cycle costs for wastewater collection and treatment; residential water reuse or recycling).
- *The possibility of separating urine from feces in toilets* (considerations include reduced sewage volumes, reduced wastewater strength/loadings including reduced loadings of pharmaceuticals, potential recovery of nutrients N and P in separated urine, reduced energy demand for aeration at WWTFs, reduced GHG loadings from WWTFs).
- *The need for research on, and subsequent adoption of, improved anaerobic processes as a pretreatment step in wastewater treatment to produce methane, to reduce organic load to aerobic processes* (considerations include recoverable energy production, reduced energy demand, reduced GHG emissions, reduced waste biological solids for treatment and disposal).
- *A critical assessment of the strategy that minimizes solids production to the exclusion of other considerations such as energy or material recovery and net environmental benefit like carbon footprint* (considerations include life cycle or triple bottom line assessment, including GHG emissions).
- *The need for research on, and subsequent adoption of, improvements in energy recovery from the liquid process trains in wastewater treatment (e.g. microbial fuel cells, hydrodynamic turbines, heat pumps) for use at wastewater treatment plants* (considerations include on-site energy production, reduced energy demand, reduced GHG emissions, life cycle assessment to determine environmental benefit).
- *The trade-off between increasingly stringent wastewater effluent quality criteria versus overall environmental quality (e.g., use of BNR vs. nitrous oxide production, need for reverse osmosis or ozone treatment of effluents vs. electricity demand and emissions; removal of metals and personal care products to biosolids vs. beneficial use).*

(considerations include energy demand, overall carbon footprint and GHG emissions, use of risk assessment in setting effluent limits relative to air emissions or biosolids disposal).

- *The need to evaluate the potential for material recovery for residual WW solids on an elemental basis (carbon, nitrogen, phosphorus, sulfur and water, at least initially)* (considerations include life cycle or carbon footprint analysis of traditional element recovery processes relative to recovery from WW solids, including energy use and GHG emissions).
- *The need to adopt a LCA or carbon footprint analysis for evaluation of biosolids disposal alternatives (land application of anaerobically digested, composted or alkaline-stabilized biosolids, landfilling, or incineration* (considerations include life cycle costs, GHG emissions, various forms of energy use, value of potential resource recovery).

Many research priorities for small and decentralized treatment facilities, as proposed by Leverenz and Tchobanoglous (2007), can be applied to addressing the points noted above. These authors have proposed four major categories of research priorities, including:

- Technologies for management of wastewater sources
- Technologies for collection, containment and transport of wastewater;
- Wastewater treatment and/or reclamation; and
- Wastewater reuse and dispersal options.

In consideration of the larger picture of the global environment, research priorities need to be extended to provide much needed data on carbon footprints and the relative sustainability of various treatment processes. Because this information is often difficult to acquire for many processes, especially with newer technologies, manufacturers of the various processes should be encouraged to provide an estimate of total life cycle costs and carbon footprint. Although some manufacturers may not wish to disclose publicly such information, it could become a marketing advantage for the more environmentally sustainable and benign technologies. Alternatively, demonstration or benchmarking projects that supply these needed data should be considered to guide municipalities in making wise longer-term decisions.

8.0 Gaps in Knowledge

8.1 Identification of Gaps

Several key data were not found during this study because the information was not available in the literature consulted. The goal of this section is to summarize and discuss the gaps in knowledge, followed by recommendations to document the required data.

The gaps identified are the following:

- Energy balance,
- Capital and O&M costs,
- Quantity of raw material used and resources produced,
- Technologies for
    - P recovery from metal precipitates,
    - coagulant recovery and recycling,

- Recovery of products on an elemental basis (e.g. nitrogen, phosphorus, carbon, sulfur, water),
- Life cycle analysis of the technologies,
- Identification of carbon footprints and GHG emissions,
- Social acceptance surveys and
- Modeling energy and resource recovery technologies and
- Optimal pathways for sludge treatment.

8.1.1 Energy Balance

The energy balance is a key parameter that can help to compare the performance of energy recovery technologies. This parameter can not be estimated without energy input and output data. Energy input and output data for some of the technologies described in the report were not available in literature. Examples of technologies without energy data reported in the literature are, Biothelys®, EBARA, Thermylis®HTB, and Bioterminator[24/85]. The output energy for the Ultrasonic sludge treatment process was also not identified.

Energy requirements for several resource recovery technologies also could not be identified. Such technologies include the Seaborne, Biocon, Sephos, Crystalactor, Phostrip and ARP processes.

Opportunities for maximizing biogas production are underway in various locations by addition of food and restaurant wastes. Research on use of the organic fraction of municipal solids waste is lacking, and could help to improve the energy balance, and reduce methane emissions from waste landfills.

8.1.2 Capital and O&M Costs

Capital and O&M costs are critical parameters for comparing the technologies and making decisions. The economic feasibility assessment is one component of the triple bottom line assessment. This costing information is needed for resource recovery technologies including Biothelys, Kopf, EBARA, Thermylis HTFB and Athos. Cost information is needed for almost all resource recovery systems. Moreover, a detailed breakdown of the costs needs to be documented. This is important because it is difficult to accurately compare the costs of different processes if the detailed breakdown is not available. Costs associated with carbon footprints are generally not recognized, but should be included in the capital cost development.

8.1.3 Quantity of Raw Material Used and Resources Produced

In addition to the sludge itself, other raw materials may be necessary for the treatment. For example, many of the technologies identified in this report use chemicals such as acids, caustic or lime. The quantity of a chemical required for the sludge treatment process, however, was not available for all of the technologies. It is important to know the quantities of all raw materials involved in each process, because it has a direct impact on the O&M costs of the processes and on a determination of like cycle assessment and social acceptance.

The quantity of the resource(s) that can be recovered from a process is also an important indicator of the performance of the process. Quantitative data was not available for all of the technologies.

8.1.4 Technologies for P Recovery from Iron Precipitates

In wastewater treatment, phosphorus can be removed as iron phosphate using $FeCl_3$ or as aluminum phosphate using alum. Ferric chloride is the most widely used coagulant in Europe for the physical-chemical removal of P, resulting in enrichment of iron phosphate in the residual sludges. The most valuable product for the phosphate industry, however, is calcium phosphate. Currently there is no commercial technology to recover phosphate from iron phosphate. Sulfate-reducing bacteria could potentially be used to release P from $FePO_4$ precipitates. The principle of this reaction is based on stronger affinity between sulfides and iron than between iron and phosphates (Jeanmaire, 2001).

8.1.5 Technologies for Coagulant Recovery and Recycling

Literature indicates that coagulants used during wastewater treatment can be recovered during sludge treatment with a process like AquaReci^{TM} (Stendahl and Jäfverström, 2004). However practical applications have not been found during this study.

8.1.6 Life Cycle Analysis

A limited number of Life Cycle Analysis (LCA) studies as been carried out in some of the European countries like Germany, France and Switzerland between 1995 and 1999 (EC 2001). Those study results were too general (including sludge spreading on agricultural lands) to be useful for this assessment, and they are now 8-12 years old. A large number of technologies has been developed or improved during the last decade. Specific LCA studies comparing energy and resource recovery technologies were not found in the literature for those technologies. In a similar vein, reporting of carbon footprints or greenhouse gas emissions resulting from implementation of the energy or resource recovery processes was generally non-existent.

8.1.7 Social Acceptance Surveys

No social acceptance studies of the technologies were found in the literature consulted. An assessment of public perception and knowledge of biosolids recycling has been conducted by Robinson and Robinson (2006) in a four county metropolitan area in the US Southeast. This study, however, did not deal with energy and resource recovery. The authors investigated eight biosolids land application site options, including farmlands, grazing lands, public parks, highway medians, home gardens, lawns, forest lands, and use of mulch after composting. It is difficult to measure the public perception of the technologies without performing surveys and statistical analyses.

8.1.8 Modeling Energy and Resource Recovery Technologies

Modeling can be a very important tool that allows prediction and/or optimization of the performance of processes. Currently, there is a little information regarding models that have been developed for energy and resource recovery from sewage sludge. Apart from the well know anaerobic digestion model No1 (ADM1), only a few models for energy or resource recovery were found during this study. Mininni et al. (1997) developed a design model of sewage sludge incineration plants to examine the possibilities for energy recovery. This model has been set-up for sizing the equipment in order to investigate different energy recovery possibilities.

Battistoni et al (2001) developed a mathematical model of struvite crystallization. After testing the model in a pilot plant they used it to design the Treviso struvite crystallization plant in Italy. Chaparro and Noguera (2002) developed a model that can be used to predict and enhance phosphorus release during anaerobic co-digestion of waste activated sludge and primary sludge. This model may help to increase struvite precipitation from the digester supernatant. A model for struvite growth for the Ostara process has been developed by a research team of the University of British Columbia, Canada (Fattah et al., 2007). None of these models are integrated into resource recovery economic models.

The performances of most the technologies reviewed in this work are based on the limited number of applications. The results may therefore not be easily applied to other sites if they are not supported by mechanistic models which are based on process operating conditions and sludge quality. Interest in modeling of energy and resource recovery from sewage sludge treatment is expected to increase rapidly.

8.1.9 Optimal Pathway for Sludge Treatment

Another gap identified is the optimal pathway assessment applied to sludge processing. Different qualities of sludge can be generated depending on the processes used for the liquid stream treatment. The optimal pathway for energy and resource recovery will depend on the quality of the feed sludge used. Considerations in determining the optimal pathway include energy potential, and concentrations of useful metals and other constituents, such as phosphorus. According to the technical literature, anaerobic digestion of raw natural primary sludge produces more biogas per mass of volatile soids (VS) destroyed than does chemically enhanced primary sludge (Dentel and Gosset, 1982; Ghyoot and Verstraete, 1997). With chemical precipitation, sludge has higher proportions of less biodegradable organic compounds due to the association of the organics with metal hydroxide flocs. Primary sludge settled without metal coagulants produces more biogas, and energy, than does excess activated sludge (Gonzàlez et al., 2003). For phosphorus recovered form sludge, an excess amount of phosphorus in the cement lowers the short-term resistance of cement (Jeanmaire, 2001). The identification and selection of optimal pathway for sludge treatment is therefore an important issue. This issue will be further addressed in Section 8.2.

8.2 Summary of the Knowledge Gaps of the Technologies

Tables 8-1 and 8-2 summarize the gaps in knowledge for each technology. An entry of "Yes" in the matrix table indicates there is a knowledge gap, while a "No" entry indicates there is no gap because the required information is available. The two tables provide an overview of the effort required to document the information not found during the study. While many of the energy and resource recovery processes have been evaluated from technical and economic positions, environmental and social acceptance reviews are lacking in virtually all cases. These considerations are incorporated into evaluations that consider the sustainability of the processes, such as a life cycle assessment or Triple Bottom Line approach.

Table 8-1. Summary of the gaps for the energy recovery technologies.

	Technology							
Process Category	**Type**	**Energy Balance**	**Capital & OM Cost**	**Quantity Raw Material & Resource**	**LCA**	**Social Survey**	**Process Modeling**	**Optimal Pathway**
Sludge-to-Biogas								
Anaerobic digestion	Bioterminator	Yes	Yes	No	Yes	Yes	Yes	Yes
Thermal hydrolysis	Cambi®	No	No	No	Yes	Yes	Yes	Yes
Thermal hydrolysis	BioThelys®	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Cell destruction	MicroSludge™	No	No	No	Yes	Yes	Yes	Yes
Cell destruction	Ultrasonic	Yes	No	Yes	Yes	Yes	Yes	Yes
Cell destruction	Ozonation	No	Yes	No	Yes	Yes	Yes	Yes
Cell destruction	Pulse Electric	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Sludge-to-Syngas								
Gasification	Kopf	No	Yes	No	Yes	Yes	Yes	Yes
Gasification	EBARA	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Incineration	Thermylis® HTFB	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Sludge-to-Oil								
Pyrolysis	EnerSludge™	No	No	No	Yes	Yes	Yes	Yes
Pyrolysis	SlurryCarb™	No	No	No	Yes	Yes	Yes	Yes
Hydrothermal	STORS	No	No	No	Yes	Yes	Yes	Yes
Sludge-to-Liquid								
SCWO	Aqua Reci®	No	No	No	Yes	Yes	Yes	Yes
SCWO	Aqua Citrox®	Yes	No	Yes	Yes	Yes	Yes	Yes
SCWO	Athos®	No	Yes	No	Yes	Yes	Yes	Yes

Table 8-2. Summary of the gaps for the resource recovery technologies.

Technology Type	Energy Balance	Capital & OM Cost	Quantity Raw Material & Resource	P recovery form iron Precipitates	Coagulant recovery	LCA	Social Survey	Process Modeling	Optimal Pathway
P from sludge									
KREPO	Yes	Yes	No	Yes	Yes	Yes	Yes	Yes	Yes
Seaborne	Yes	Yes	Yes	*Not applicable*	Yes	Yes	Yes	Yes	Yes
Aqua-Reci®	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Kemicond™	Yes	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes
P from Ash									
BioCon	Yes	Yes	Yes	*Not applicable*	Yes	Yes	Yes	Yes	Yes
SEPHOS	Yes	Yes	Yes	*Not applicable*	Yes	Yes	Yes	Yes	Yes
P from Side-stream									
Crystalactor®	Yes	Yes	Yes	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
Phostrip©	Yes	Yes	Yes	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
Ostara	Yes	Yes	Yes	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
Nitrogen Recovery									
ARP	Yes	Yes	Yes	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
Building Material									
TS[a]-ALWA	Yes	Yes	No	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
TS[a]-Slag	Yes	Yes	No	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
TS[a]-Brick	Yes	Yes	No	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes
GlassPack®	No	No	No	*Not applicable*	*Not applicable*	Yes	Yes	Yes	Yes

[a] Thermal solidification

8.3 Recommendations Resulting from Knowledge Gap Analysis

Recommendation 1

An improved economic evaluation of the recovery processes requires additional data regarding the costs, energy balance of the technologies and amount of resources that can be recovered. This type of information can not be obtained without the participation of the owners and users of the technologies.

Recommendation 2

To compare the environmental impacts of different sludge treatment technologies, use of LCA is necessary. These LCA studies will reinforce the third component of TBL assessment. As a measure of environment performance, LCA gives both quantitative and qualitative measures. Inclusion of carbon foot printing to the Capital planning processes would be useful.

LCA studies are particularly necessary for the future and emerging technologies. Such studies will require that the input and output gaps be determined in order to increase the accuracy of the results. For example, a LCA study of the Seaborne process can not be accurately addressed without knowing the quantity of energy and chemical used for the treatment. Energy credit for a reduction of fossil fuel use can not be assigned to a process like Bioterminator[24/85] if the amount of net biogas energy generated is not known.

Recommendation 3

A new framework is needed for assessing wastewater treatment (including but not limited to solids processes) based on elemental (C, N, P, S) pathways and cradle-to-grave optimization of the overall net environmental benefits. The new assessment framework should identify pathways to most efficiently recover energy, elements, and water throughout the treatment train when not constrained by current infrastructure configurations. Emphasis of the assessment would include a focus on processes which create new products for reuse, which examine the trade-offs between maximizing one factor over another (e.g. sludge volume reduction versus increased CO_2 production potential from emerging processes), and which maximize resource recovery and net environmental benefit.

Recommendation 4

In order to perform a better evaluation of the second component of the TBL assessment (social acceptance), a survey must be designed and administered to residents in different selected countries members of the GWRC. The survey will invite the residents to give their opinion regarding each of the technologies described in the report. The result of the survey will then be treated statistically, and procedures that assign values to social considerations can be applied.

Recommendation 5

The use of sulfate-reducing bacteria to release of P from $FePO_4$ precipitates should be further investigated. Effort should focus on reaction rate improvement and commercial technology development. If such technology were developed and proved to be efficient, it may result in a broader application (beyond P recovery from sludge) because iron phosphate exists to a greater extent in nature than as calcium phosphate ore. The economic feasibility of the process should also be assessed.

Recommendation 6

The technical and economic feasibility of coagulant recovery from sewage sludge needs to further demonstrated. The efficiency of the recovered coagulant when recycled should be also addressed.

Recommendation 7

Research that promotes more efficient and cost effective clean up of biogas (such as the removal of siloxanes, the recovery of elemental sulfur, and the removal of carbon dioxide from biogas) is needed to improve the cost effectiveness of biogas for combined heat and power recovery.

Recommendation 8

Research on co-digestion of the organic fraction of municipal solid waste (otherwise landfilled or incinerated) is required to maximize the production of biogas from the anaerobic digestion of wastewater solids. Use of other organic wastes (i.e. fog, food wastes) to improve biogas production when co-digested with wastewater solids is established or in current research projects.

GLOSSARY

Carbon Footprint (CF): Measures the amount of CO_2 or CO_2 equivalents emitted by an activity.

Biogas: refers to the gas produced by the fermentation of organic matter including, sewage sludge, under anaerobic conditions. Biogas is comprised primarily of methane and carbon dioxide.

Biosolids: Solids generated from the treat of sewage sludge with processes such as anaerobic digestion, aerobic digestion, lime stabilization, etc.

Char: Carbon-rich substance produced in absence of oxygen gasification or pyrolysis of the volatile solids in sludge.

E-fuel: Carbonized product produced during the treatment of the sludge with SlurryCarb^{TM} Technology.

Energy: The term energy used in this report include electricity, heat and fuel.

Greenhouse Gases (GHGs): Gases in the atmosphere that contribute to Global Warming. Those gases include but are not limited to, carbon dioxide, methane and nitrous oxides.

Global Warming (GW): Global warming is an increase in the near surface temperature of the Earth.

Life Cycle Analysis (LCA): Analysis that examines the impact (energy use, greenhouse gases and air pollutants) a product has on the environment from the beginning to the end of its lifetime.

Recycling: The re-processing of materials into new products. May also apply to the return of material from one process or location to an upstream process or location.

Resources: Materials or products other than energy that can be recovered from sludge such as phosphorus, building material, etc.

Sewage Sludge: Excess solids produced in municipal wastewater treatment plants.

Sludge-to-Biogas Processes: Processes that convert sludge into biogas or enhance the conversion of sludge to biogas.

Sludge-to-Oil Processes: Processes that convert sludge to oil.

Sludge-to-Syngas Processes: Processes that convert sludge to syngas such as gasification.

Sludge-to-Liquid: Processes that convert sludge to liquid such as super critical oxidation.

Surface Water: A stream, river, lake or wetland. May also include any water overlying land above sea level.

Syngas: Gas produced during gasification or pyrolysis of sewage sludge. Syngas gas typically contains nitrogen (N_2), carbon dioxide (CO_2) and carbon monoxide (CO) and hydrogen (H_2).

Triple Bottom Line (TBL): TBL captures an expanded spectrum of values and criteria for measuring organizational (and societal) success; economic, environmental and social.

Volatile Acids: Low molecular weigh organic acids, such as acetic, propionic and butyric acids) produced during the initial stage of anaerobic biodegradation of sludges. Also called volatile fatty acids.

Volatile Fatty Acids: (see Volatile Acids).

ABBREVIATIONS

ARP	Ammonia Recovery Process
BNR	Biological Nutrient Removal
BTU	British Thermal Unit
COD	Chemical Oxygen Demand
DS	Dry solids
ESI	Environmental Solutions International
EU	European Union
FBF	Fluidized Bed Furnace
GBP	British Pounds (currency)
GHG	Greenhouse gas
GWRC	Global Water Research Coalition
HGG	Hot Gas Generator
ITC-WGT	Institute for Technical Chemistry - Water and Geo-Technology Division
Kemicond	Kemira sludge Conditioning
KREPO	Kemwater REcycling PROcess
KTH	Kungl Tekniska Högskolan (Royal Institute of Technology)
LCA	Life Cycle Analysis
LCAMER	Life Cycle Assessment Manager for Energy Recovery
LPG	Liquefied Petroleum Gas
LSEC Center	Lake Shinji East Clean Center
MAP	Magnesium Ammonium Phosphate (struvite)
MT	Metric Tonne
MHF	Multiple Hearth Furnace
MWTPs	Municipal Wastewater Treatment Plants
NCG	Non-Condensed Gas
P	Phosphate or phosphorus
PAC	Powder Activated Carbon
P.E	Population Equivalent
P-RoC	Phosphorus RecOvery by Crystallization
PS	Primary Sludge
ROCK	Review of Current Knowledge
RW	Reaction Water
SAS	Surplus Activated Sludge

SCWO	Super Critical Water Oxidation
SoS	State of Science
STORS	Sludge-To-Oil Reactor System
TBL	Triple Bottom Line
TS	Total Solids
TVS	Total Volatile Solids
VS	Volatile Solids
UK	United Kingdom
US	United States
U.S. EPA	United States Environmental Protection Agency
WAS	Waste Activated Sludge
WERF	Water Environmental Research Federation
WWTF	Wastewater Treatment Facility
WWTP	Wastewater Treatment Plant

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Appendix A:

Literature Review of Anaerobic Digestion and Energy Recovery from Wastewater Sludge

Introduction to the Appendix

This Appendix is a review of the technical literature that encompasses anaerobic sludge digestion technology, digester gas (biogas) pretreatment for downstream energy recovery, and the energy recovery processes themselves.

This Appendix was prepared as part of the Final Report prepared by Hydromantis Inc. and the University of Toronto for the Water Environment Research Foundation under WERF Project 01-CTS-18UR. The Final Report is titled "*An Assessment Tool for Managing Cost-effective Energy Recovery from Anaerobically Digested Wastewater Solids*". The spreadsheet model Life Cycle Assessment Manager for Energy Recovery (LCAMER) resulted from this project. The inclusion of the literature review chapter from Project 01-CTS-18UR as an appendix to this report has been authorized by WERF.

Anaerobic Digester Processes

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 product of sludge digestion, mainly consisting 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 to 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:

$$C_nH_aO_b + (n - a/4 - b/2)\ H_2O \rightarrow (n/2 - a/8 + b/4)\ CO_2 + (n/2 - a/8 - b/4)\ CH_4$$

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 word. Some of these new systems are the upflow anaerobic sludge blanket (UASB) rector (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).

Anaerobic Digestion Process Configurations

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 A-1 shows the schematic of a low-rate digester.

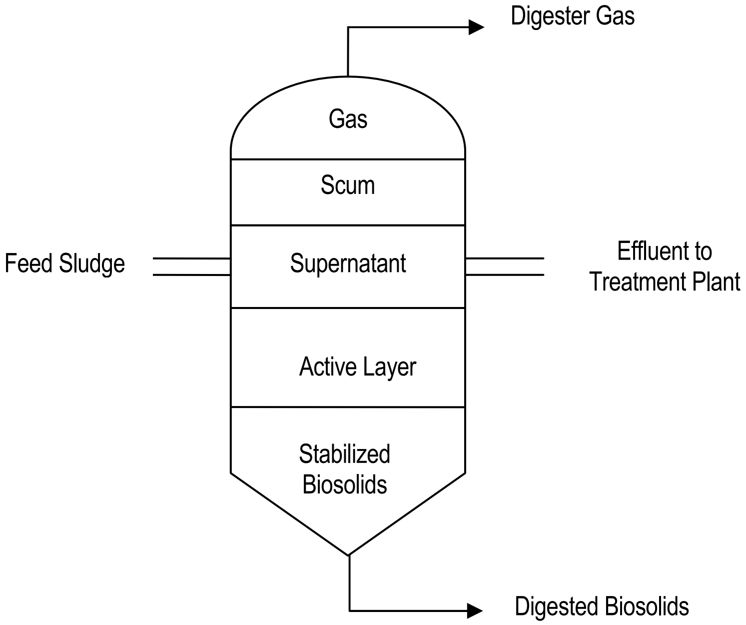

Figure A-1. Low-rate Anaerobic Digestion.

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 3450m^3/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).

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 A-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^{o}C (86-100^{o}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 m^3 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 A-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 m^3/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).

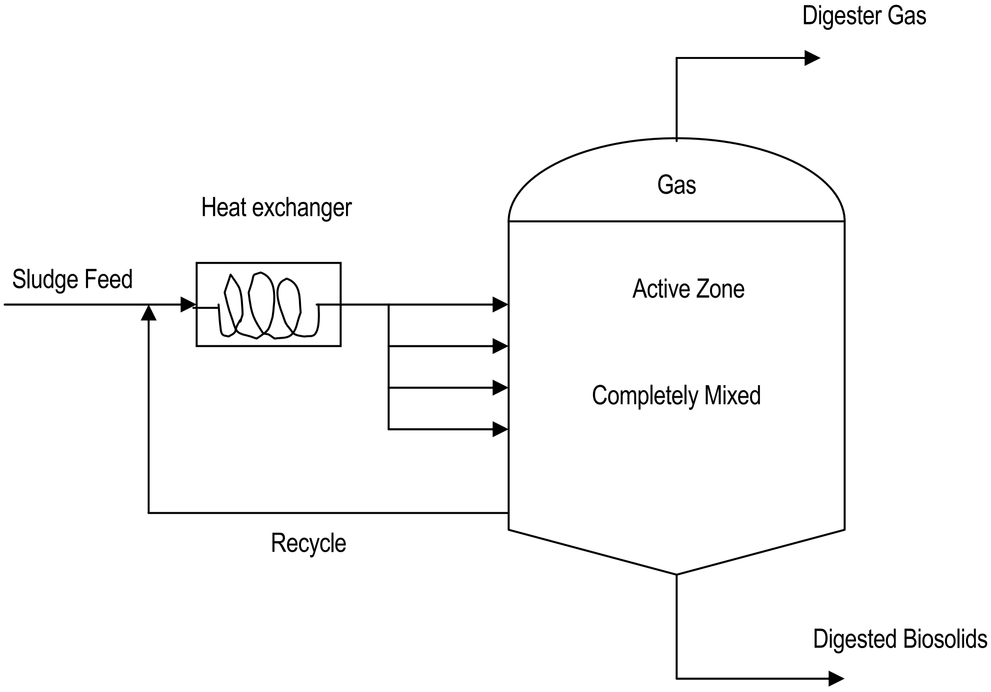

Figure A-2. High-rate Anaerobic Digestion.

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 C_5 and C_6 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 A-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.

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 A-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.

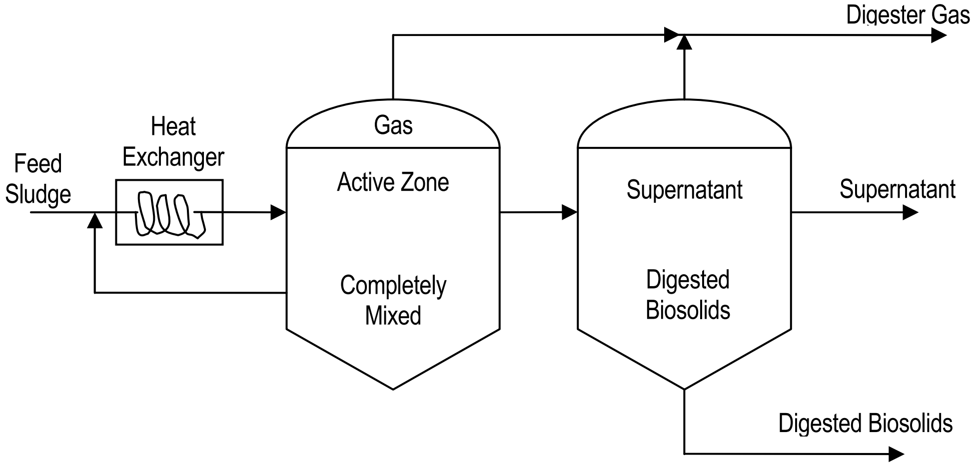

Figure A-3. Two-stage Anaerobic Digestion (unmixed in 2nd stage).

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 A-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.

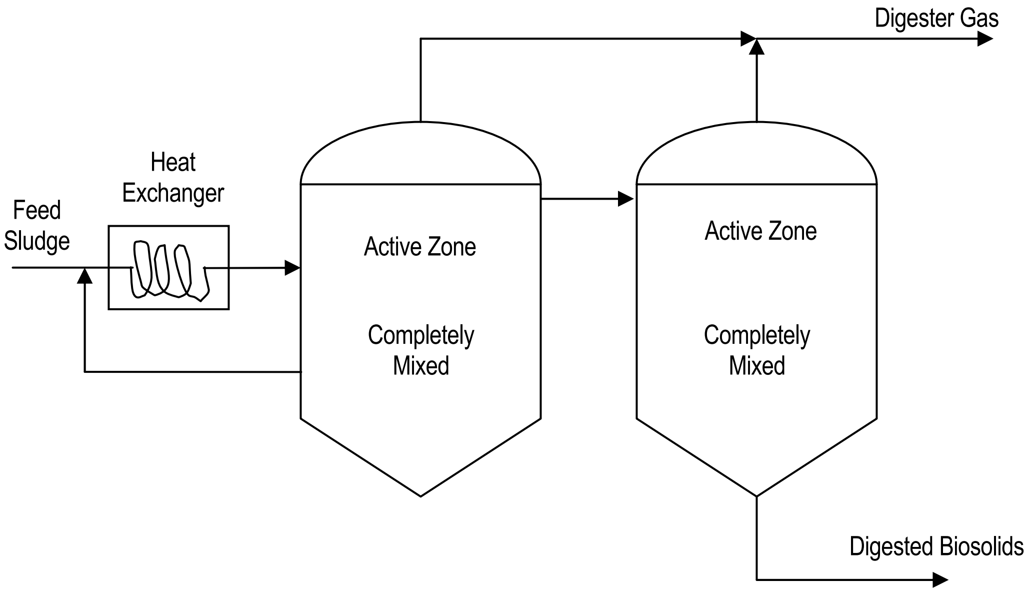

Figure A-4. Two-stage Anaerobic Digestion (mixed in 2nd stage).

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).

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) (FigureA-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).

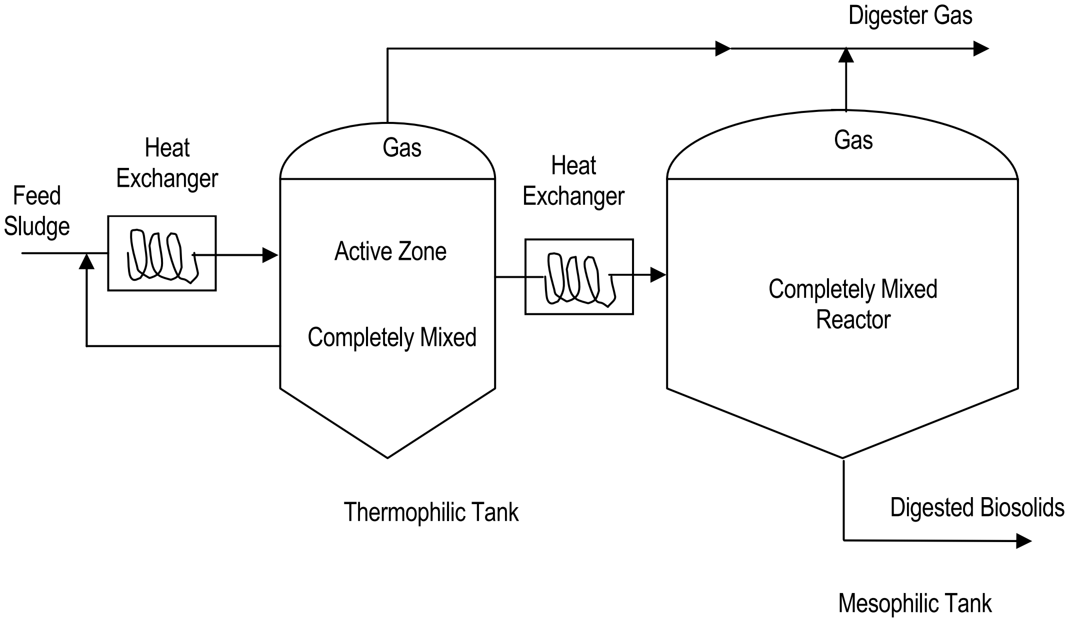

Figure A-5. Temperature-phased Anaerobic (Thermo+Meso) (TPAD).

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;
- 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 U.S., 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 A-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).

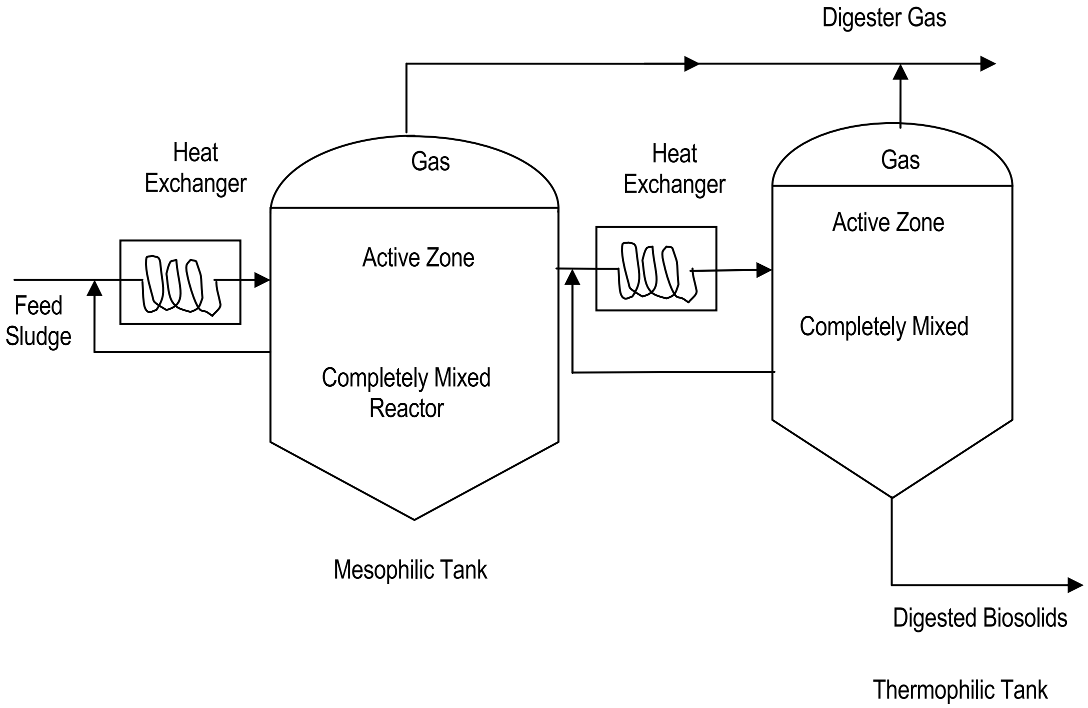

Figure A-6. Temperature-phased Anaerobic Digestion System (Meso+Thermo) (TPAD).

There are also other modes of TPAD. A three-stage TPAD (Figure A-7) is used in the 227,000 m^3/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).

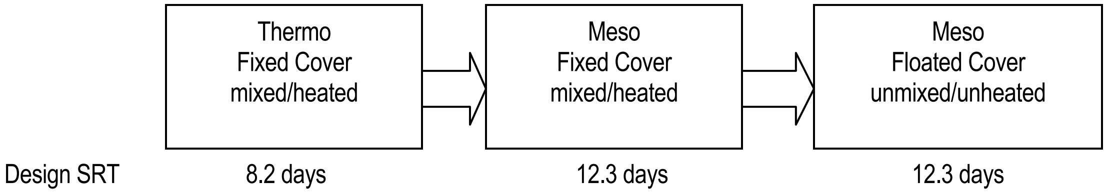

Figure A-7. Design of temperature-phased anaerobic digestion in Village Creek WWTP.

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 A-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).

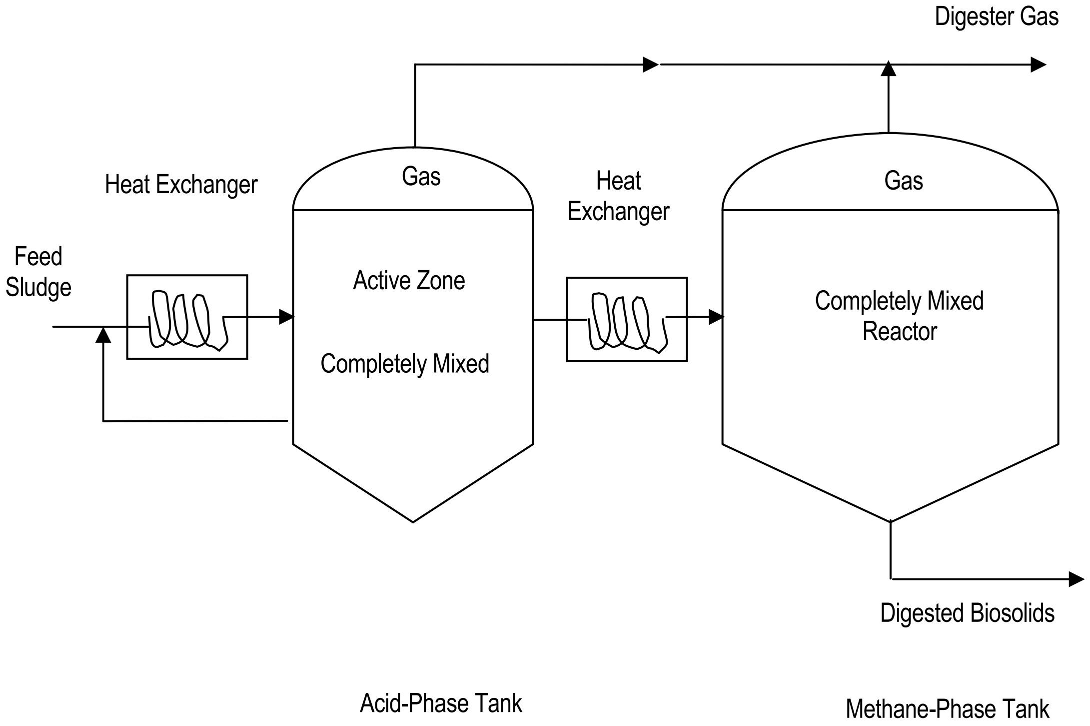

Figure A-8. Two-phase Anaerobic Digestion (Acid-gas Phased Anaerobic Digestion).

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 H_2 and CO_2 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.

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 A-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).

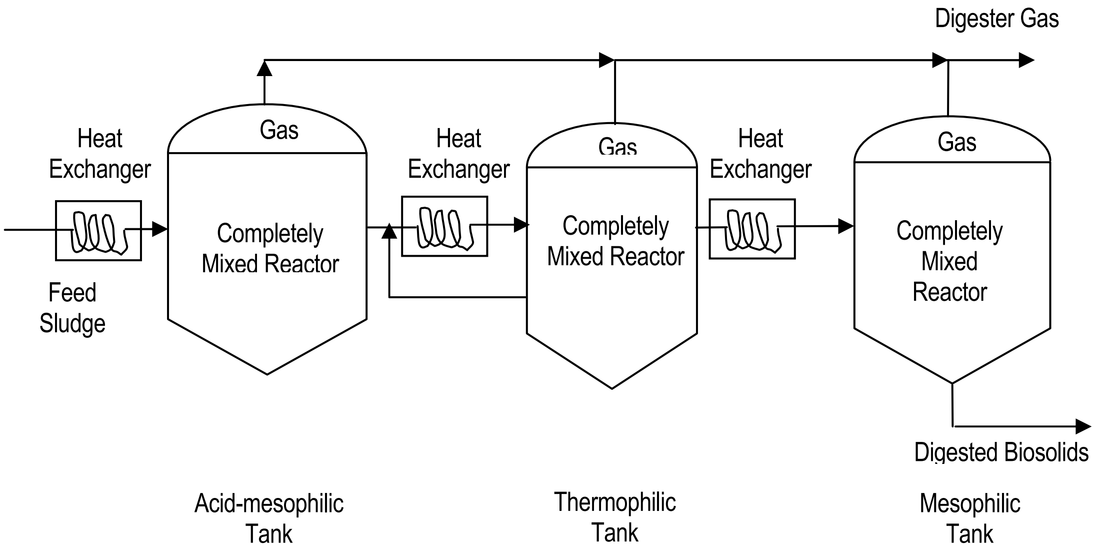

Figure A-9. Combination of 2-phase and TPAD Anaerobic Digestion System.

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).

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 A-1.

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

Technology	Advantages	Disadvantages
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 Amaerica 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 al., 1998). The results of a survey in Germany about egg-shaped digesters showed that mixing energies for egg-shaped digesters are typically only 40% to 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 10 times (or more) greater than through a concrete one.

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 (BOD_5) 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 BOD_5 and 100- 10,000 mg/L (0.0008- 0.084 lb/gal) for TSS. Total BOD_5 concentration in digester supernatant is the sum of soluble BOD_5 and BOD_5 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 BOD_5 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.

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 (H_2O), CO_2, hydrogen sulfide (H_2S) (Constant et al., 1989) and siloxanes (Tower, 2003b) are known to have corrosive effects on ERS.

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 vapour 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 vapour removal efficiencies so cannot be the only form of moisture removal for ERSs sensitive to water vapour such as microturbines.

Desiccant dryers have higher water vapour 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 and 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).

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.

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).

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/m^3 (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/MJ$_{fuel}$ (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).

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 245,000m^3/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 ERSs in Europe and U.S. SELEXOLO manufactured by Union Carbide is a popular type, and is capable of removing H_2S and CO_2, 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, U.S.) was about US$ 50,000 (US$ 8.20 per m^3/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 m^3/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 A-2 details other plants that used activated carbon to pretreat digester gas.

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

Table A-2. Plants using Activated Carbon for preatreating digester gas.

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

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 H_2S (Osinga, 2000).

Impact on Energy Recovery Systems and the Environment

The presence of H_2S 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 H_2S. Also, when water vapour is present during combustion, sulfuric acid (H_2SO_4) 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 H_2S is particularly harmful to internal combustion engines. However, the presence of H_2S 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_2S has on ERS performance, its main combustion product, sulfur dioxide (SO_2) reduces ambient air quality significantly (Constant et al., 1989).

Hydrogen Sulfide Removal Techniques

Technologies developed for removing H_2S from natural gas and refinery gas are fairly well established. However, because digester gas has higher levels of H_2S than natural gas and refinery gas, these processes have limited application with digester gas (Osinga, 2000).

A summary of the various H_2S removal processes is presented in Table A-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_2S.

Table A-3. Summary of The Various H_2S removal Techniques[1].

Techniques	H_2S Inlet Concentrations	Level of H_2S Treatment	Waste Product	Capital Cost (US$)	Annual Operation & Maintenance Cost (US$)
Activated Carbon	Low	>85%	Spent media	10.38/ m^3/day(2) (0.294/scf/d)	0.004 /m^3/day(2) (0.00011/scf/d)
Iron Salts Addition	>6,000 ppmv	>85%	Iron sulfide	0.05/ m^3/d (0.0002/gpd) capacity flow (3)	0.35/m^3/d (0.0013/gpd) capacity flow (3)
Iron Sponge	200-5,000ppmv	90%	Spent iron sponge (containing iron oxide, sulfur, soda, ash, limestone, wood)	81.00/ m^3/d (2.29/scf/d) digester gas flow	12/ m^3/d (0.034/scf/d) digester gas flow
SulfaTreat	<500ppmv	>85%	Iron disulfide	9.5/ m^3/d (0.269/scf/d)digester gas flow(4)	17.5/ m^3/d (0.496/scf/d) digester gas flow (4)
BINAX		>99%	Water saturated with HS^-	187/m^3/d (5.295/scf/d) digester gas flow (5)	Not Available
LO-CAT	30-8,000ppmv	>98%	Elemental sulfur	Not Available	6.40/ m^3/d (0.181/scf/d) digester gas flow
Apollo	200-20,000ppmv	>99%	Elemental sulfur	Not Available	6,390/yr

(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_2S 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_2S to elemental sulfur, which is deposited on the carbon particles. However, this alternative is only suitable for small volumes of gas and low H_2S concentrations (Earth Tech (Canada) Inc. et al., 2002).

As mentioned in Table A-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_2S, 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_2S from digester gas. It is currently licensed by Barnebey and 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_2S 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 A-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:

$$Fe^{2+} + 2Fe^{3+} + 4HS^{-} \rightarrow Fe_3S_4 + 4H^{+} \qquad (3\text{-}1)$$

This method removes H_2S 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, smythite, 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 A-4, the dosage and the removal effectiveness differs between WWTPs. Generally, levels below 200ppm of H_2S 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). Accodring to (Dezham et al., 1988), the annual chemical cost for adding $FeCl_2$ to reduce H_2S 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_2S 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).

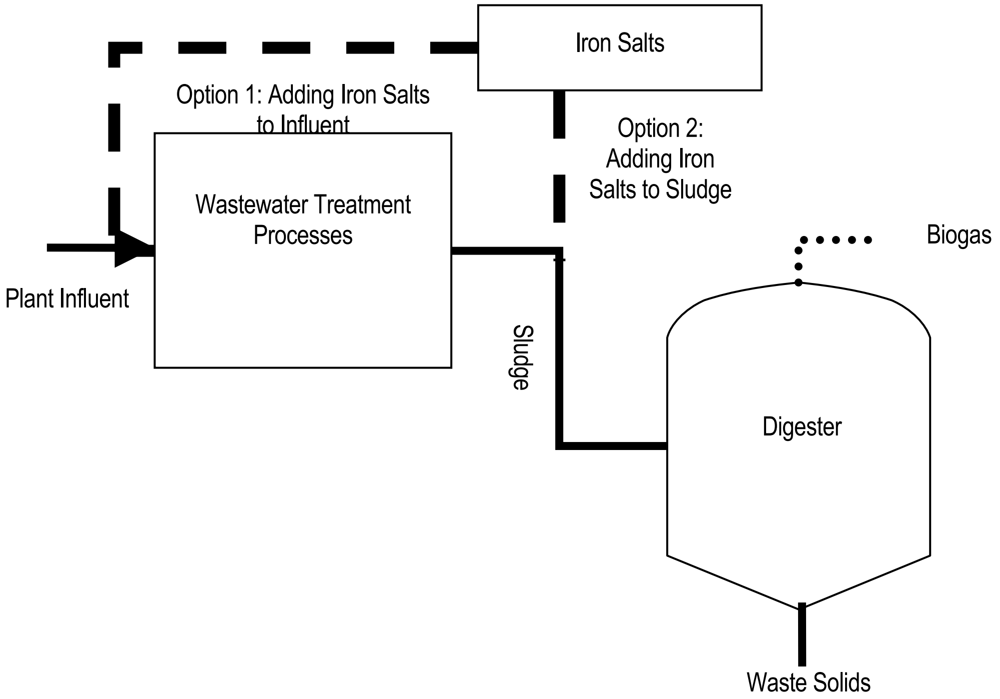

Figure A-10 Schematic Representation of an Iron Salt Process,

Table A-4. Iron Chloride Dosages for H_2S reduction(1).

Plant	**H_2S reduction**	**Dosage**	**Other factors**
City of Santa Maria WWTP, CA	2900 to 950 ppm	16kg $FeCl_2$/1000 kg VSS (16 lb $FeCl_2$/1000 lbVSS)	Further reduction to 300ppm with 3.9mg/l (0.0002 lb/scf) $FeCl_2$ or 4.4mg/l (0.0003 lb/scf) of FeCl3 added
City of Lethbridge WWTP, Alberta	2400ppm to 600ppm	600 L/d (21.2 scf/d) 37% FeCl3	HRT reduced from 40 to 25 days decreased H_2S 6800ppm to 2400ppm
San Jose/Santa Clara WPCP, CA	300ppm final concentration	5-9.5 kg (11-30 lb) $FeCl_2$ or 15-27 kg (33-60lb) FeCl3	
Encina WPCF, Carlsbad, CA	From >1000ppm to 200-300ppm		Continuous chemical feed reduced demand for the ferric compared to batch feeding
Sacramento, CA (Kido et al, 1995)	1500ppm to 165ppm		Added iron chloride to headworks.
Longmount WWTP, Colorado (Bielefeldt et al, 2002)	52-95% reduction for 1.2 – 74 ppm conc	2 mg/L (0.0001 lb/scf) ferric chloride to headworks (42% solution - ~33gpd into ~9mgd flow)	

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 H_2S 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 ($Fe_3PO_4 \cdot 8H_2O$) 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 BOD_5, TSS, or nitrogen would not be significant (Dezham et al., 1988).

The relative cost of iron salt pretreatment in comparison to other available H_2S 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 H_2S. It has been found to be cost effective for lowering H_2S 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 H_2S from digester gas. It is capable of reducing the H_2S 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, U.S.), and the JWPCP are some of the WWTPs that are using this alternative to control H_2S 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 ($2Fe_2O_3 \cdot H_2O$). A fiberglass grate supports the media (Kido et al., 1995). The chemical reaction of the process is as follows:

$$2Fe_2O_3 + 6H_2S \rightarrow 2Fe_2S_3\ (s) + 6H_2O \qquad (3\text{-}2)$$

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

Low-pressure gas flows through the dry media, and the H_2S in the influent gas stream combines with the iron oxide to form iron sulfide (Fe_2S_3) and water (Earth Tech et al., 2002; Kido et al., 1995; Ravishanker and Hills, 1984). Gas should be wet when passing through the

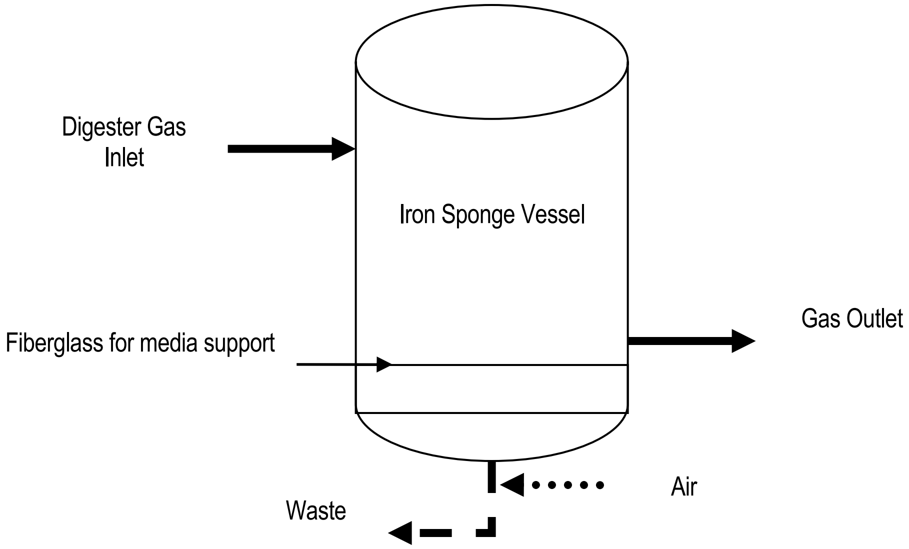

Figure A-11 Schematic Representation of an Iron Sponge Vessel

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):

$$2Fe_2S_3 + 3O_2 \leftrightarrow 2Fe_2O_3 + 6S_{(s)} \qquad (3\text{-}3)$$

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 H_2S removal and allows for a higher gas flow rate (Earth Tech et al., 2002). In parallel iron sponge operation, the inlet H_2S 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 m^3/day (145,295-1,056,688 gpd). Varec units are reported to reduce H_2S 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 H_2S 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 3,870m^3/d (1,022,346 gpd) and an inlet H_2S concentration of 1500 ppmv, the installed capital cost of an iron sponge system is approximately US\$ 312,500 (US\$ 81/m^3/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/(m^3/d digester gas)/yr or US\$ 0.340 /(scf digester gas)/yr) (Earth Tech et al., 2002). Table A-5 lists some WWTPs that use an iron sponge.

Table A-5. Wastewater Treatment Plants that use Iron Sponge Pretreatment.

Plant	H_2S reduction	Dosage
Union Sanitary District WWTP (Earth Tech *et al.*, 2002)	200 to 50ppm	8 units
Central Marin Sanitation Agency WWTP, San Rafaell, CA (Kido et al, 1995)	400 to 35 ppm	2 units in series (replaced media 3 times in 8 years)

SulfaTreat: The SulfaTreat system (Figure A-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 H_2S, and last two to three times longer (NATCO Group, 2002). The SulfaTreat system has been used since 1998, mainly in the natural Figure A-12 Schematic Representation of a SulfaTreat System 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 – 30,000ppm, and are purified to non-detectable limits (Al-Issa, 2004). H_2S removal to less than 1ppmv 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).

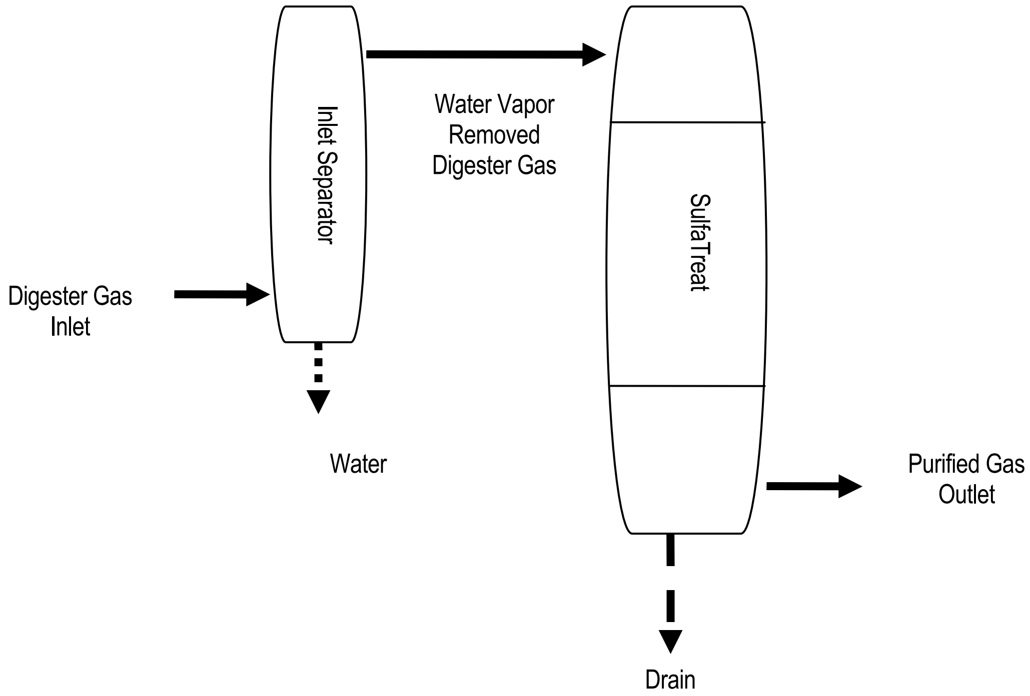

Figure A-12 Schematic Representation of a SulfaTreat System.

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°C to 60°C (40° to 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 H_2S concentrations would typically be operated in series. The amount of media required (hence the vessel size) would be determined by the H_2S 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 H_2S 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 H_2S 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 (EarthTech 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 0.5tpd, (Earth Tech et al., 2002).

The City of Lakeland WWTP (Florida, U.S.) and the Stickney Water Reclamation Plant, in Chicago, (Illinois, U.S.) 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/m^3/d digester gas flow (US$ 0.27 scf/d digester gas flow) and the associated annual operating cost is approximately US$ 17.50/ (m^3/d digester gas flow)/year or US$ 0.50/ (scf/d digester gas flow)/y for an average digester gas flow rate of 3870 m^3/d (1,022,000 gpd) with an average H_2S concentration of 1500 ppmv (Earth Tech et al., 2002).

Water Scrubbers: Water-only scrubbers operate on the principle of H_2S solubility in water (Osinga, 2000). H_2S absorbed in the water dissociates into HS- and S_2- ions. Chemical equilibrium depends on the H_2S 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 CO_2 and H_2S 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 CO_2 to 4-5% (Constant et al., 1989). A generic water scrubber can take in 200- 2500 ppm H_2S (Earth Tech et al., 2002). Removal of H_2S 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 H_2S and CO_2 (details on CO_2 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 H_2S removal only, and the maximum is 300psig, which is required to remove both H_2S (to less than 1ppm), and CO_2 (Osinga, 2000).

Figure A-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 CO_2 and H_2S 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 H_2S is reduced to 4 ppmv in a BINAX system (Henrich, 1984).

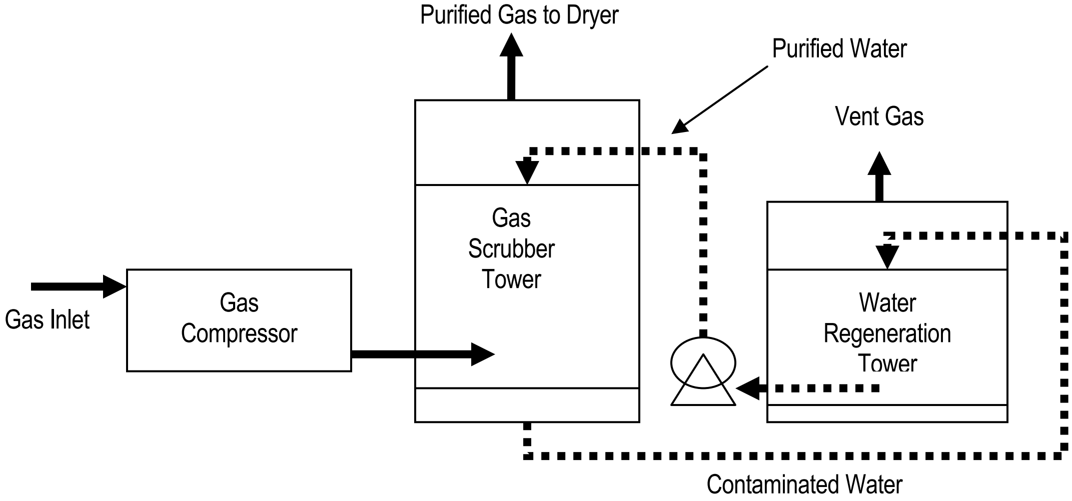

Figure A-13 Schematic Representation of a BINAX System.

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 2,680m^3/d (94,643 gpd) system (US$ 187/m^3/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 m^3/day (1.2M ft^3/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 425-42500m^3/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 H_2S 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 H_2S to elemental sulfur, and one in the oxidizer section, which regenerates the catalyst (see Figure A-14). The absorber reactions are as follows (Gas Technology Products LLC, 2003):

$H_2S_{(g)} \leftrightarrow H_2S_{(aq)}$ (3-4)

$H_2S_{(aq)} \leftrightarrow HS^- + H^+$ (3-5)

$HS^- + 2Fe^{3+} \rightarrow S_{(s)} + 2Fe^{2+} + H^+$ (3-6)

In the absorber, H_2S is absorbed into the slightly alkaline, aqueous LO-CAT solution. The H_2S 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):

$O_{2(g)} \leftrightarrow O_{2(aq)}$ (3-7)

$O_{2(aq)} + 4Fe^{2+} + 2H_2O \rightarrow 4Fe^{3+} + 4OH^-$ (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):

$H_2S_{(g)} + \frac{1}{2}O_{2(g)} \rightarrow H_2O + S^0$ (3-9)

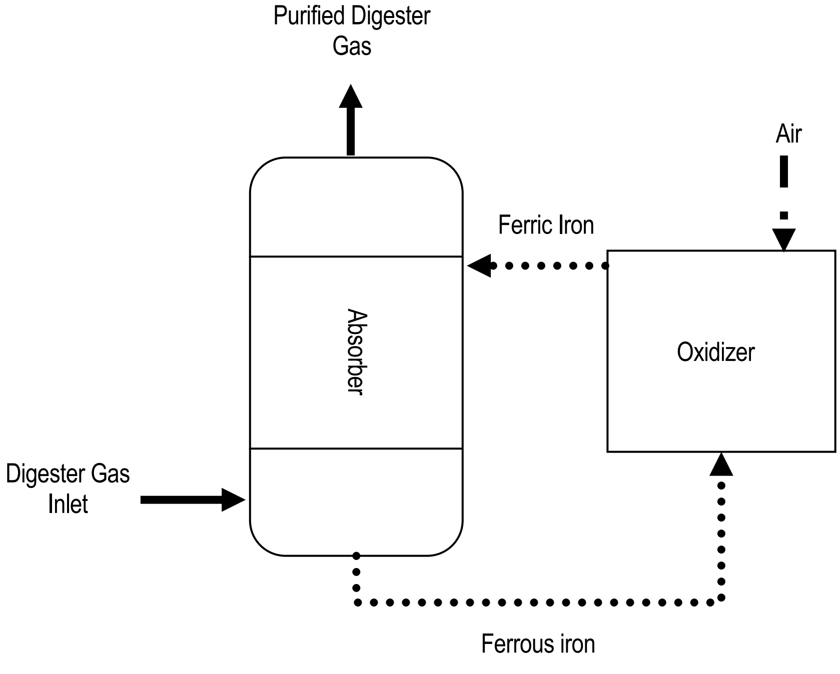

Figure A-14 Schematic Representation of LO-CAT System.

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^{o}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 H_2S 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, U.S.) has reported chemical consumption costs of US$ 25,000/yr or US$ 6.40/(m^3/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 H_2S, which are outlined in Table A-6 (Gas Technology Products LLC, 2003; Katehis et al., 2003; Earth Tech et al., 2002).

Table A-6. Information on Lo-Cat Hydrogen Sulfide Removal Efficiency.

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

Earth Tech et al, 2002

Apollo: The Apollo gas scrubber removes sulfides and particulates from anaerobic digester gas that contains H_2S concentrations up to 20,000 ppmv, at a flow rate of 3,900 to 193,000m^3/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 H_2S 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

$$H_2S_{(g)} \leftrightarrow H_2S_{(aq)} \qquad (3\text{-}10)$$

Ionization

$$H_2S_{(aq)} + OH^- \leftrightarrow H_2O + HS^- \qquad (3\text{-}11)$$

Substitution

$$HS^- + FeOHEDTA^{2-}{}_{(aq)} \leftrightarrow FeHSEDTA^{2-}_{(aq)} + OH^- \qquad (3\text{-}12)$$

Oxidation

$$FeHSEDTA^{2-}_{(aq)} + FeOHEDTA^{2-}_{(aq)} \leftrightarrow 2FeEDTA^{2-}_{(aq)} + S^0 + H_2O \qquad (3\text{-}13)$$

Regeneration

$$H_2O + \tfrac{1}{2}O_{2(aq)} + 2FeEDTA^{2-}_{(aq)} \leftrightarrow 2FeOHEDTA^{2-}_{(aq)} \qquad (3\text{-}14)$$

Overall Reaction

$$H_2S + \tfrac{1}{2}O_2 \leftrightarrow H_2O + S^0 \qquad (3\text{-}15)$$

The system is not affected by fluctuating flowrates or H_2S 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 twenty times faster than conventional liquid contactors (OCETA, 2002b).

A schematic representation of the Apollo system is presented in Figure A-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_2S 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_2S removal efficiency of 98% for 100-1000ppmv H_2S (Earth Tech et al., 2002).

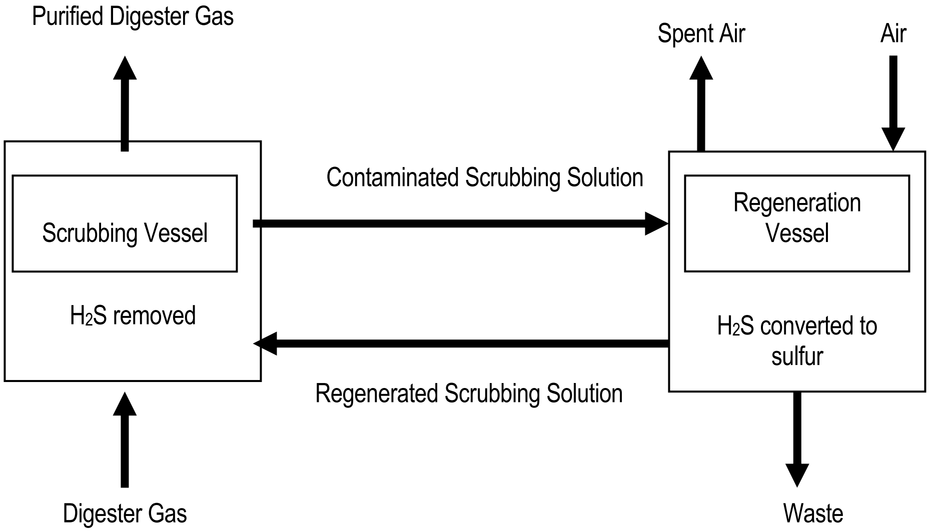

Figure A-15 Schematic Representation of an Apollo Scrubber Unit and Recirculation System.

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, U.S.) 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).

Digester Gas Energy Recovery Systems

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 A-16.

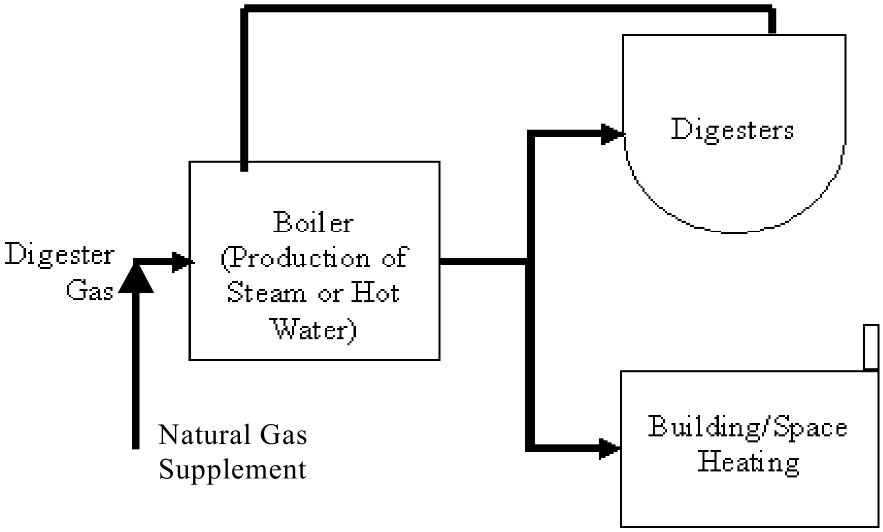

Figure A-16. Schematic of Energy Recovery from Digester Gas using Boilers

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, H_2S 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 CO_2 and NO_x.

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

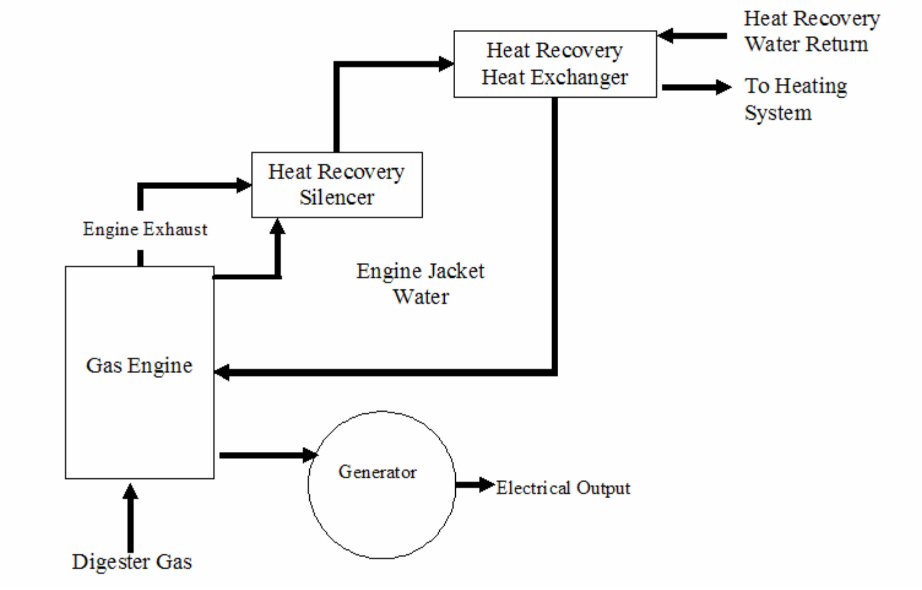

Figure A-17. Schematic of Energy Recovery from Digester Gas using Engine Generators

Microturbines

Microturbines are a newer ERS that can be used in designated ozone Non-Attainment Areas because of the low NO_x 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^{o}C (65^{o}F). With combined heat and power (CHP) recovery, the overall efficiency can rise to the order of 70 to 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 A-18. Identified microturbine installations at POTWs are summarized in Table A-7.

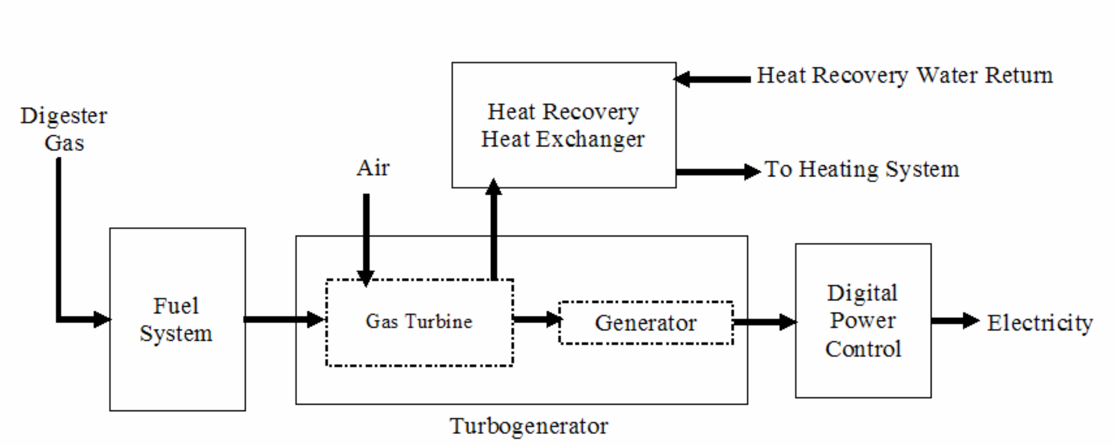

Figure A-18. Schematic Diagram of Energy Recovery from Digester Gas from a Microturbine Unit.

Table A-7. Summary of Microturbine Cogeneration in North America (Chiu, 2004).

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

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 NO_x and SO_x, as well as low emissions of CO_2. A number of different types of fuel cells are available, operating at different temperatures and using different electrolytes, as indicated in Table A-8 (Chiu, 2004).

Table A-8. Types of Fuel Cells.

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

[1] Source: Appleby, 1996
[2] Source: Hordeski, 2003

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

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

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

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

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 A-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, U.S.) and at Los Angeles' Terminal Island facility (California, U.S.), uses molten carbonate technology for their stationary fuel cells, with power outputs ranging from 250 kW to 2000 kW. Energy recovery efficiency ranges from 47-50%. 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, U.S.), 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 A-11 while operating and technical information for the fuel cell types are provided in Table A-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 A-19.

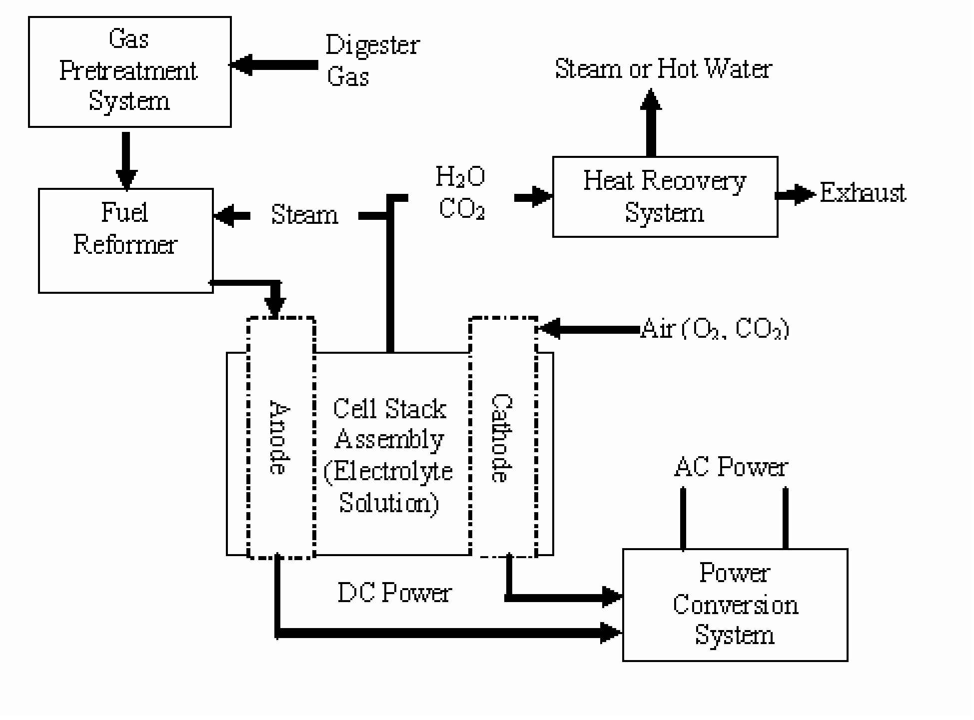

Figure A-19. General Schematic of Fuel Cell Cogeneration System.

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 A-13.

Table A-10. List of Suppliers of Fuel Cells for Wastewater Treatment Plants.

Name of the Fuel Cell Manufacturer			Name of the contact person					
	Product(s)	**Website**	**Last Name**	**First Name**	**Title**	**Phone**	**Fax**	**Email**
TESTING								
Fuel Cell Energy Inc.	DFC	http://www.fce.com	Baker	Bill	Contacts for News Media	(860)350-9100	N/A	N/A
COMMERCIALLY AVAILABLE								
UTC Fuel Cells	PC25 (200kW)	Http://www.utcfuelcells.com	Frankhauser	Greg	Regional Sales Manager	(410)740-5616	N/A	Greg.Frankhauser@UTCFuelCells.com
PLANNED FUTURE PROGRAM								
Fuel Cell Technologies LTD.	Modular 50kW commercial unit	Http://www.fct.ca	Allen Adams	Gary Mike	Sales (SOFC) Sales (Al/O_2)	(613)544-8222	(613)544-5150	gallen@fct.ca madams@fct.ca

Table A-11. List of Fuel Cell Installations.

Fuel Cell Manufacturer	Project Partner	Fuel Cell	Location	Building	Status	Fuel used	Cost of Project(USD)	Comments
Fuel Cell Energy Inc.	King County, Washington	1MW DFC®	King County, Washington	Municipal WWTP	Construction-current (April 2003); Operation(October,2003 expected)	Digester Gas	$1.8-2.2 million USD depending on source	
Fuel Cell Energy Inc.	LADWP [1]	250kW DFC300®	Los Angeles, California	Fuel Cell Power Plant at Terminal Island in San Pedro	To be delivered	Digester Gas from a nearby WWTP	Unknown	Project follows after LADWP's 250kW DFC300® in Los Angeles(March2003)
UTC Fuel Cells	City of Portland, Oregon	200kW PAFC(PC25)	Portland, Oregon	Columbia Blvd. WWTP	In Operation since May 1999	Digester Gas	~$1.3 million USD	
UTC Fuel Cells	Energy 2000 Las Virgenes Municipal Water District and Truunfo Sanitation District	Two 200kW PAFC (PC25)	Calabasas, CA	Rancho Las Virgenes Composting Facility	In Operation since 1999	Digester Gas from nearby WWTP	~$2.6 million USD	
UTC Fuel Cells	NYPA[2],KeySpan With Westchester County WWTP in Yonkers, NY	200kW PAFC PC25	Yonkers, NY	Westchester County WWTP	In Operation since April 1997	Digester Gas		Supplies Supplemental Power, Grid Parallel, World's First ADG fueled fuel cell
UTC Fuel Cells	NYPA and NYSERDA [3]	Three 200kW PAFC(PC25)	Coster Street and Ryawa Ave, Bronx,NY 10474	Hunts Point WWTP	Under Construction (2003)	Digester Gas	N/A	
UTC Fuel Cells	NYPA and NYSERDA	Two 200kW PAFC (PC25)	63 Flushing Ave. Brooklyn, NY 11205	Red Hook WWTP	Under Construction (2003)	Digester Gas	N/A	
UTC Fuel Cells	NYPA and NYSERDA	Two 200kW PAFC(PC25)	43-01 Berrian Blvd., Astoria, NY 11105	Bowery Bay WWTP	Under Construction (2003)	Digester Gas	N/A	
UTC Fuel Cells	NYPA and NYSERDA	One 200kW PAFC(PC25)	751 Mill Rd., Staten Island, NY 10306	Oakwood Beach WWTP	Under Construction (2003)	Digester Gas	N/A	

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[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 A-12. Operating and Technical Information on Fuel Cells Systems.

Company Name	Product	Types	Market Entry Date	Fuel Used	Operating Temperature	Power Output	Efficiency	Waste Heat/Emission	Comment
Fuel Cell Energy Inc.	DFC 300 Stationary Fuel Cell Unit	Molten Carbonate	Testing	Fuel Flexible		250kW	47%	From typical natural gas: NO_x<0.3ppmv SO_x<0.1ppmv CO<10ppmv Exhaust Temperature:650˚F Exhaust Heat Available:300,000BTU/hr	
Fuel Cell Energy Inc.	DFC 3000 Stationary Fuel Cell Unit	Molten Carbonate	Testing	Fuel Flexible		2000kW	50%	From typical natural gas: NO_x<0.3ppmv SO_x<0.01ppmv CO<10ppmv Exhaust Temperature:~650˚F Exhaust Heat Available:2.8 mm BTU/hr	
Fuel Cell Energy Inc.	DFC 1500 Stationary Fuel Cell Unit	Molten Carbonate	Testing	Fuel Flexible		1000kW	49%	From typical natural gas: NO_x<0.3ppmv SO_x<0.01ppmv CO<10ppmv Exhaust Temperature:~650˚F Exhaust Heat Available:1.4mm BTU/hr	
Fuel Cell Technologies LTD.	Modular 50kW Commercial Unit	Tubular SOFC	Planned Future Program	Fuel Flexible for first generation. Second generation will use diesel fuel and furnace oil	700 to 1000˚C (1290-1830)	5kW	~90% with co-generation	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
UTC Fuel Cells	PC25 200kW PAFC Systems	PAFC	Available	Natural Gas, Digester Gas or both		200kW	87% with co-generation	On 15% O_2 dry basis: NO_x<1ppmv SO_x<negligible CO<2ppmv	

Table A-13. List of Wastewater Treatment Plants Using Direct Drive Cogeneration System.

Plant	City	Type of Energy Recovery System
Valley Creek	Birmingham, AL	Engine Blowers
North Regional	Dayton, OH	Engine Generators and Engine Pump
Northside	Durham, NC	Engine Blowers
Village Creek	Forth Worth, TX	Engine Generators and Engine Blowers
Nansemond	Virginia Beach, VA	Boilers and Engine Blowers
Muddy Creek	Winston-Salem, NC	Boilers and Engine Blowers

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/m^3/d (US\$82,000/MGD) plant capacity flow, and the annual operating cost is estimated to be about US\$1/(m^3/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 A-20.

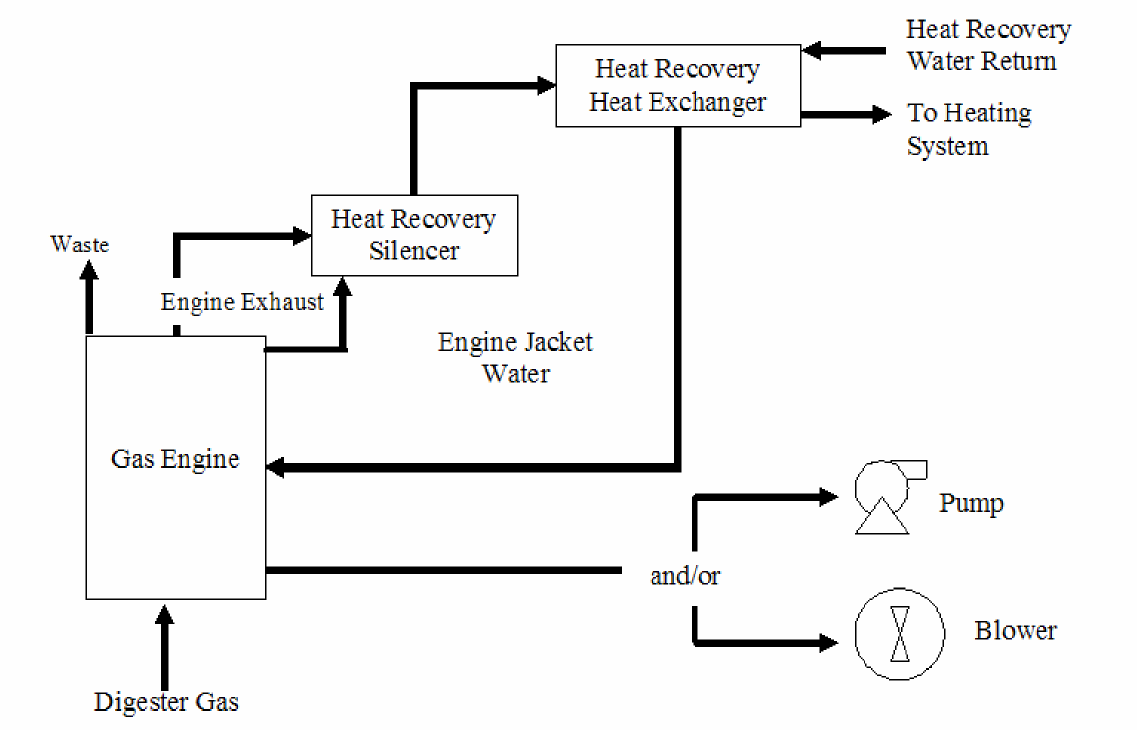

Figure A-20. Schematic Diagram of Energy Recovery from Digester Gas Using Direct Drive Equipment.

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 $\$0.008/m^3$ ($30/million gallons) treated, compared to $\$0.009/m^3$ ($34/million gallons) and $\$0.012/m^3$ ($45/million gallons) for the electrical energy generation and heat generation schemes, respectively.

Stirling Engine

Stirling engines operating on digester gas are of interest mainly because they can operate without costly GPS for H_2S 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 A-14.

Table A-14. Advantages and Disadvantages of Stirling Engine System.

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

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 H_2S 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, U.S.) although it was not operated with cogeneration. The technology has been tested to withstand H_2S 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^{o}C (20^{o}F) below the gas temperature. The range for the gas temperature is 43-60^{o}C (110-140^{o}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).

Summary of Energy Recovery Systems

Pertinent information regarding ERSs are summarized in Table A-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 A-15. Summary of the Various Energy Recovery Systems.

ERS	Gas Pretreatment Requirements[1]	Thermal Efficiency	Electrical Efficiency	Capital Cost[2]	O&M Costs[2]
Boiler	None	75-80%	N/A	\$100/kWh ($23,500/hp)	\$0.56/MJ ($470/hp-yr)
Gas Engine Cogeneration	A,B,C	45-50%	30-35%	\$1,000/kW-\$1,250/kW	\$13/kW/yr
Gas Engine Direct Drive	A,B,C	~46%	26-28%	**\$22/m3/d** ($82,036/MGD)	\$1.10/m3/d ($4,219/MGD)
Microturbine	A,C	35-40%	24-28%	\$1450/kW	\$130/kW/yr
Phosphoric Acid Fuel Cells (PAFC)	A,B	~37%	~40%	\$6250/kW	\$250/kW/yr

[1] A = Water Vapor; B = H_2S; C = Siloxane

[2] Capital and operating and maintenance costs are reported in U.S. Dollars (USD)

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Appendix B:

Detailed Descriptions of Existing Energy and Resource Recovery Technologies

B.1 TECHNOLOGIES FOR PHOSPHORUS RECOVERY

B.1.1 Crystalactor® Technology

Description

The Crystalactor® is a vertical cylindrical fluidized bed reactor which combines coagulation, flocculation, separation and dewatering in a single reactor (DVH, 2007). The reactor is filled with sand as seeding material. Crystallisation of calcium phosphate takes place mainly onto the surfaces of the sand. With time, the calcium pellets increase in size and weight. Larger and denser pellets move to the bottom of the reactor. The pellets are discharged from the bottom of the reactor and new sand material is added. The pellets typically consist of 40-50% calcium phosphate, 30-40% sand, and up to 10% calcium carbonate (STOWA, 2006a).

A solution of lime [$Ca(OH)_2$] is added to the reactor to increase both the pH, to about 8, and the concentration of calcium ions. This allows optimizing the calcium phosphate precipitation. The crystallisation efficiency can also be enhanced by recirculation. With a recirculation ratio of 2.5 to 3 an efficiency of approximately 70% can be achieved (STOWA, 2006a).

The Crystalactor® reactor is usually incorporated as one unit process within a biological nutrient removal (BNR) system. Part of the sludge from the secondary clarifier of the BNR system is pumped to an anaerobic phosphate stripping tank in the side-stream. Under the anaerobic conditions, ortho-phosphate is released from the sludge to the liquid phase. The sludge and the phosphate-rich water are separated. The sludge is recycled to the to BNR system while the phosphate-rich water is fed to a cascade stripper to remove carbon dioxide (CO_2). Thereafter, the phosphate-rich water, containing 50 to 80 mg P/L, flows through the Crystalactor reactor. CO_2 stripping is necessary to prevent the formation of calcium carbonate in the Crystalactor reactor.

Application

The Crystalactor® Technology has been applied at full-scale in the Netherlands. The technology was installed at the Geestmertambacht, Heemsted and Westerbrork (Stack, 2007). The Geestmertambacht experience is described below.

In 1994, the Geestmertambacht plant had a capacity of 230,000 P.E (STOWA, 2006a). The Crystalactor® reactor was installed to recover phosphorus as calcium phosphate which could be used as raw material in the phosphate industry (e.g. Themophos B.V.) to produce phosphoric acid. The capacity of the installation is 250 m^3/h, producing 70 kg phosphate/h (DVH, 2007).

Cost Estimate

The total investment cost, as of 1996, for the side-stream process was estimated at about €4.2 million (about US$6.3 million). The capital cost was estimated to be about 55% of the total annual cost. This was equivalent to €7300 (about US$10950) per metric tonne

(MT) of P removed (Gaastra et al. 1998). The cost of the process was estimated as 22 times higher than the cost of mined phosphate rock (Roeleved et al., 2004), and as a consequence, the Geestmerambacht MWTP installation was judged economically poor (Roeleved et al., 2004). However, there are hopes that with the gathered experience simpler and more effective design of the reactor could be implemented, resulting in savings of capital costs of about 5 - 10% (STOWA, 2006a).

B.1.2 Technical Variants of Crystalactor®

Technical Variants

A process using tobermorite-rich waste materials from the construction industry, as seed material, is under development the Institute for Technical Chemistry - Water and Geo-Technology Division (ITC-WGT) in Germany. This process named P-ROC or Phosphorus RecOvery by Crystallization has been described by Berg and Shaum (2005). Tobermorite appears to simulate the precipitation of calcium phosphate, while it also increases the pH due it chemical properties. At the same time it serves as crystallization nucleus.

Other variants of Crystalactor have been applied at full-scale at Lake Shinji East Clean (LSEC) Center in Japan (Ueno and Fujii, 2001) and at Treviso Sewage Works in Italy (Battistoni, et al. 2001). In the LSEC Center, struvite (magnesium ammonium phosphate or MAP) was recovered from anaerobic digester supernatant and sold (Ueno and Fujii, 2001). The LSEC Center experience is further described below.

Example of Lake Shinji East Clean Center, Japan

- ***Plant Description***

LSEC Center had a treatment capacity of 45,000 m^3 per day in April 1994. The plant was designed to remove nitrogen in a two-stage BNR process. A pre-stage anaerobic and aerobic activated sludge process was used to remove phosphorous. At the same time powdered activated carbon (PAC) was added to the aeration tank. The excess sludge from the process was anaerobically digested to release phosphorus. After dewatering the digested sludge, the filtrate rich in phosphorus (about 70% of the influent P load) was returned to the treatment system. Poly-ferric sulfate and a large amount of PAC were initially used in order to achieve the required effluent P concentration.

The struvite precipitation process was installed at LSEC Center to 1) lower the effluent phosphorous concentration 2) reduce the amount of PAC and poly-ferric sulfate used and 3) produce a fertilizer (struvite).

- ***Process***

The sludge process capacity was 1,150 m^3 per day in September 2000. The diagram of the struvite precipitation process is presented in Figure B-1. The process consists of:

- A vertical cylindrical reaction tower including a separation zone at the top of the reactor,
- A cylindrical rotation separator,
- An electrically opening and shutting square cut gate hopper.

The filtrate from the sludge treatment facility is charged at the bottom of a precipitation reaction tower. The concentration of phosphoric acid in the filtrate is 100 to 110 mg/L. Magnesium hydroxide is added to the reactor. Sodium hydroxide is also injected to adjust the pH between 8.2 and 8.8. Struvite grow in the reactor with a granular shape, 0.5 to 1.0 mm after about 10 days.

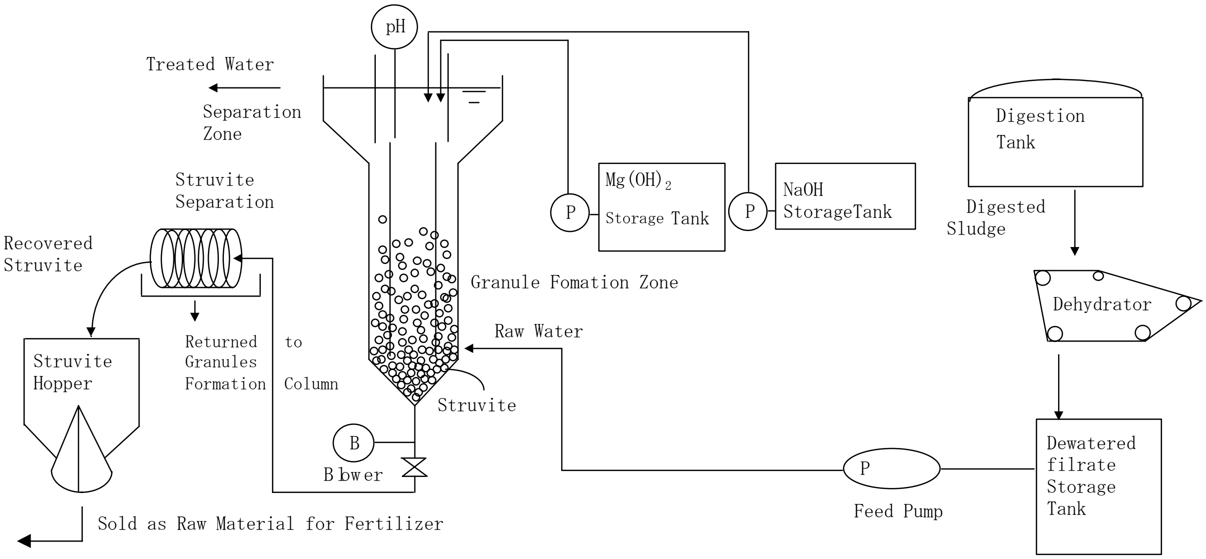

Figure B-1. Process diagram of the Struvite precipitation process used at the LSEC Center (Ueno & Fujii, 2001).

- ***Performance***

The produced struvite is recovered at the bottom of reactor and sold as fertilizer. About 500 to 550 kg of struvite is produced per day and sold at €250 per MT (US$ 375 per MT). The treated effluent leaves the reactor from the top with a phosphoric acid concentration of 10 mg/L. This process therefore achieves more than 90% removal. Table B-1 summarizes the composition of the struvite produced.

Table B-1. Composition of Struvite produced at the LSEC Center (Ueno & Fujii, 2001).

Parameters	Value
Ammonia nitrogen	5.67% N
Citric soluble phosphoric acid	29.53% P_2O_5
Citric soluble magnesia	16.18% MgO
Cadmium	0.000006% Cd
Arsine	0.000048% As

B.1.3 PhoStrip© Technology

Description

The PhoStrip© Technology is based on recovery of phosphorus from a wastewater side stream. Phosphorus is recovered from the return sludge flow in a stripper. Figure B.2 shows the simplified flow diagram of the Phostrip© Technology. This figure was initially published by Drnevich, (1979) as reported by Levlin and Hultman (2003).

A portion of the return sludge from the clarifier is pumped to an anaerobic stripper tank. Phosphorus is then released from the sludge to the liquid phase. In order to increase the amount of phosphorus released, acetic acid may be dosed to the stripper. The phosphorus-containing water is separated from the sludge. The liquid is treated with lime to precipitate the phosphorus as calcium phosphate.

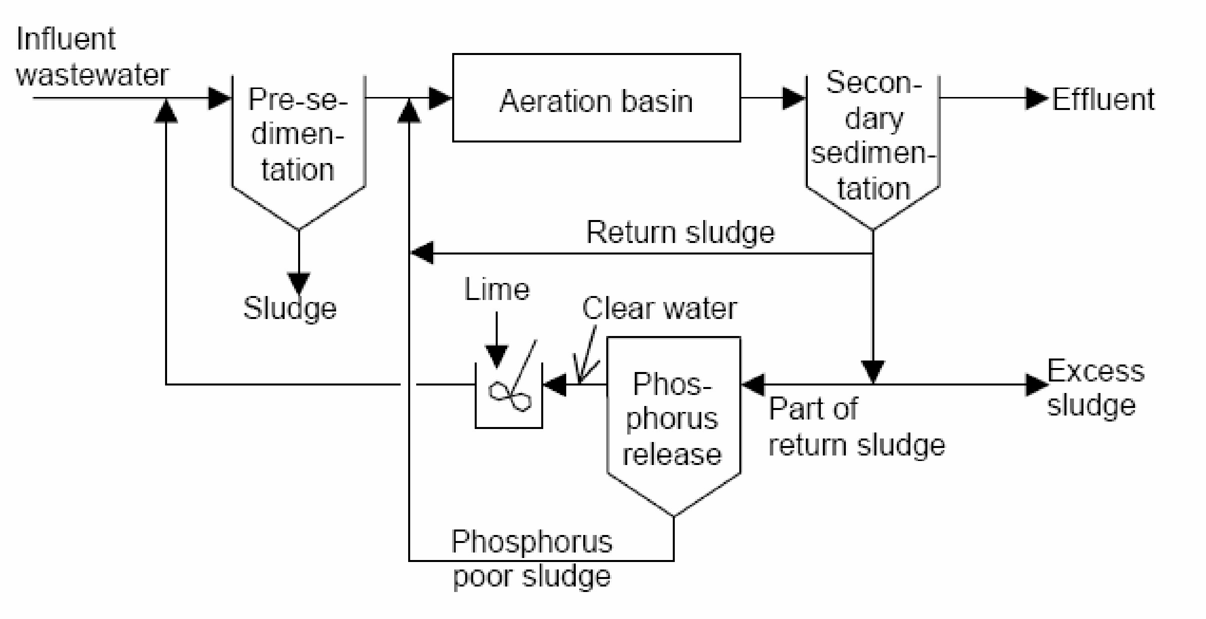

Figure B-2. Simplified flow diagram of the PhoStrip© Technology (From Levlin & Hultman, 2003).

Application
PhoStrip© was installed in the 1970s in the Reno/Sparks Joint Water Pollution Control Plant located in Sparks, Nevada, U.S. (Levlin and Hultman, 2003). The goal of this first demonstration phase of the technology was not to recover phosphorus but just to produce an effluent with low phosphorus concentration. Therefore the lime precipitated sludge was recycled to primary sedimentation tank as indicated in Figure B.2.

Since then, PhoStrip© Technology has been used in many MWTPs. At least 4 four full-scale plants were using the process in the middle of the 1980s including Lansdale, PA; Little Patuxent, MD; Central Contra Costa Sanitation District, CA and Seneca Falls, NY. The design parameters of those plants were summarized by Rybicki (1997).

B.2 TECHNOLOGIES FOR BUILDING MATERIAL RECOVERY

Typical building materials that can be produced from sewage sludge include artificial lightweight aggregates (ALWA), brick, slag, ceramic, cement, glass, interlocking tile, etc. In general, the materials are produced after incinerating the sludge. However, for brick and cement production, the sludge can be used directly without incineration. The main process used to produce the material is thermal solidification. This process consists in melting and then solidifying the ash.

Processes used to produce ALWA, brick and slag have been described by Okuno et al. (2004). A brief summary of each process is presented below. This is followed by the description of a few examples of full-scale plants in Japan.

A process developed in North America by Minergy Corporation and called GlassPack® Technology is further described in Chapter 6.0 as a case study.

B.2.1 Thermal Solidification for ALWA Production

ALWA are produced by blending ash with water (23% wt/wt) and a small amount of alcohol-distillation waste. The mixture is supplied to a centrifugal pelletizer. The pellets produced are dried at 270°C for 7-10 minutes. The dried pellets are finally heated at 1050°C in a fluidized bed kiln for a few minutes to produce the final material.

B.2.2 Thermal Solidification for Slag Production

Slag is produced by melting ash into a cyclone furnace pre-heated to 1500°C. The ash is initially blended with a small amount of lime. Slag is a marble-like mineral of semi-crystalline structure.

B.2.3 Thermal Solidification for Brick Production

The production of sludge-based bricks requires pre-treatment of the sludge ash at high pressure. This is followed by heat treatment at a temperature up to 1000°C. The material from these first two steps is finally poured into a die and pressed up to 1000 kg/cm^2.

B.2.4 Examples of Thermal Solidification Plants in Japan

Table B-2 prepared from information reported by Okuno and Yamada (2000) and Onaka (2000) shows three plants and the processes used for producing building materials with sewage sludge in Japan.

Thermal solidification processes are energy intensive. Energy consumption for slag, brick and ALWA production are summarized in Table B.3. Data reported in the last column of the report was calculated assuming 30% dry solids, on average, in the dewatered cake (Onaka, 2000).

Table B-2. Example of plants producing building materials from sewage sludge in Japan.

Plant Name	Start-up	Unit Process	Product Name	Product Quantity	Products Use
Nambu [a]	1996	-Water/alcohol Blender -Centrifugal Pelletizer -Dryer(270°C) -Fluidized kiln(1,050°C)	ALWA	500 kg per h	-Thermal insulator -Water-infiltration plate -Fillers
Minami [b]	1990	-Steam dryer -Hot blast crusher -Melting furnace (1,500°C)	Slag	540 kg per h	-Concrete aggregates -Interlocking tiles -Water permeable tiles
Tagawa [b]	1998	-Pelletizing -Drying	Pellets	210 kg per h	-Portland cement

[a] Ash from sludge incineration is used as feed, [b] Dewatered sludge cake is used as feed

Table B-3. Energy consumption for Slag, Brick and ALWA production (Adapted after Okuno & Yamada, 2000).

Product	Natural Gas (m^3/d MT cake)	Electricity (kWh/MT cake)	Total Energy [b] (kWh/MT cake)	Total Energy [c] (kWh/dry MT cake)
Slag	35.0	130	497.5	1,658
Brick-incineration ash [a]	40.4	206	630.2	2,101
ALWA [a]	38.1	156.5	556.7	1,856

[a] Include energy used during incineration step; [b] Sum of electricity and natural gas expressed in kWh; [c] assume 30% dry solids in dewatered sludge

B.3 CATEGORIES OF TREATMENT PROCESSES FOR ENERGY RECOVERY

Physical, mechanical, biological and chemical processes can be used to produce or contribute to energy recovery from sludge. These processes can be grouped into four main categories; Sludge-to-Biogas processes, Sludge-to-Syngas processes, Sludge-to-Oil processes and Sludge-to-Liquid. Examples of technologies (established and emerging) for each category are summarized in Table B-4. This section discusses only the established technologies. The emerging technologies will be discussed in Appendix C.

Table B-4. Categories of treatment processes for energy recovery.

Processes	Example of Technology
Sludge-to-Biogas	
Anaerobic digestion	Bioterminator$^{24/85}$
Thermal hydrolysis	Cambi®, BioThelys®,
Physical-chemical Cell destruction	MicroSludge™, Ultrasonic, Ozonation, Pulse electric
Sludge-to-Syngas	
Gasification	KOPF, EBARA
Incineration	Thermylis® HTFB
Sludge-to-Oil	
Pyrolysis	Enersludge™, SlurryCarb™
Hydrothermal	STORS
Sludge-to-Liquid[a]	
SCWO	Aqua Reci®, Aqua citrox®, Athos®

[a]Sludge is converted to a liquid form and heat is recovered

B.3.1 Sludge-to-Biogas Processes

Anaerobic digestion has been the traditional biochemical process used to produce biogas from sewage sludge. Many anaerobic digestion technologies have been developed. Among those technologies are, one-stage mesophilic, two-stage mesophilic, acid-gas phase, one-stage thermophilic, two-stage thermophilic, three-stage and temperature phased. All these technologies have been described in detail by Monteith et al. (2006), and are provided in Appendix A.

Much research has focused, in recent years, on increasing the biogas production during anaerobic digestion of sewage sludge. The main processes investigated to date are thermal hydrolysis and physical-chemical cell destruction. This has led to the development of different technologies including hydrothermal heating, ultrasonic cell disintegration, use of ozone and electrical pulses, etc.

Once biogas has been produced, the energy in the gas can be recovered as thermal electrical or mechanical energy. Table B-5 lists the methods for digester gas energy recovery. A more detailed discussion is provided in Appendix A.

Table B-5. Digester Gas Energy Recovery Technologies.

Processes	Example of Technology
Thermal energy	
Hot water or low pressure steam	Boilers
Combined heat and power (as electricity)	
Internal combustion engine	engine-generators
Turbines	micro-turbines, combined cycle turbines
Fuel Cells	low temperature (phosphoric acid, proton exchange membrane, alkaline); high temperature (molten carbonate, solid oxide)
Combined heat and power (as mechanical)	
Internal combustion engine	direct drive engines
External combustion	Stirling engine

B.3.1.1 Thermal Hydrolysis

Thermal hydrolysis is a process used to increase sludge digestibility and decrease the amount residue for disposal. During thermal treatment, sludge is heated at high temperature and high vapor pressure for several minutes. Microbial cell structures in the sludge are destroyed and easily digestible organic compounds are released from the destroyed cells.

The advantages of combining thermal hydrolysis and anaerobic digestion are high volatile solids (VS) destruction and increased biogas production. Currently, the most known commercial thermal hydrolysis technologies are Cambi® and BioThelys®.

B.3.1.1.1 Cambi® Technology

Description

Cambi® Technology was developed and patented by the Norwegian Company Cambi. This technology consists of a series of reactors including a pulping vessel, a hydrolysis reactor and a flash tank (Steve and Panter, 2002). All three reactors are closed and the system operates batch-wise. The Cambi® system can be used to treat both undigested primary and secondary sludge.

Figure B-3 shows the flow diagram of a typical Cambi® Technology. Solids are first dewatered to approximately 16% dry solids. The dewatered sludge is transferred to the pulping vessel where it is pre-heated to approximately 80°C by steam addition. The pre-heated sludge is then sent to the thermal hydrolysis reactor. This reactor operates at a temperature of 160°C to 180°C and a pressure of 600 kPa (6 bars) for about 30 minutes (Potts and Jolly, 2007). After the thermal treatment, the sludge is transferred to the flash tank operating at atmospheric pressure. The change in pressure from the hydrolysis reactor to flash tank causes cell lysis. The hydrolyzed sludge is then cooled for mesophilic anaerobic digestion.

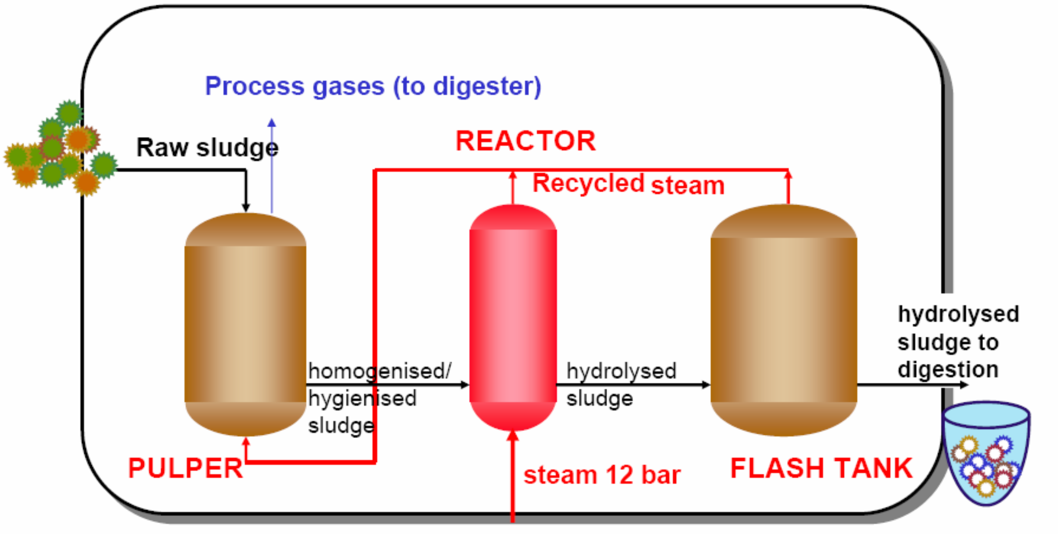

Figure B-3. Process flow diagram of a typical Cambi® Technology (From Fjaergard et al. 2006).

Possible issues with the Cambi® process can include odour problems (Steve and Panter 2002), mechanical problems due to fibrous material in primary sludge, and chemical attack of the cavity pump used to transfer sludge into the reactors and the digester (Pickworth et al., 2005). Steve and Panter (2002) reported that thermal hydrolysis produces about 30 m^3 gas per hour with a very strong odour. All the above problems are avoided if the installation is properly designed and appropriate equipments are used (Steve and Panter 2002; Pickworth et al., 2005).

Examples of Full-scale Applications – HIAS MWTP, Norway

Since 1995 at least ten full-scale Cambi® installations have been built in different countries. These installations operate in the range of 1000 to 36000 dry MT of sludge per year (Panter and Kleiven, 2005). Table B-6 summarizes some of the existing installations operating in different countries, all located in Europe. No operating installations were identified in North America.

Table B-6. Cambi® installations in different countries (Pickword et al., 2005a; Keep et al., 1999; Panter & Kleiven, 2005).

MWTP Name	Location	Capacity (dry MT per year) [a]	Start-up
HIAS MWTP	Norway	3600	1995
Chertsey MWTP	UK	8000	1998
Naestved MWTP	Denmark	1600	2000
Borregaard MWTP	Norway	4000	2000
Nigg Bay MWTP	Scotland	15000	2001
Ringsend MWTP	Ireland	36000	2002
Frederica MWTP	Denmark	8000	2002
Kapusciska MWTP	Poland	7500	2005

[a] Capacity of the installation

- ***HIAS MWTP Plant Capacity***

HIAS MWTP in Hamar, Norway serves approximately 50,000 P.E., although during high organic loading periods the waste load to the plant corresponds to approximately 125,000 P.E. The plant is designed for removal of organic matter and phosphorus in three steps (Kepp et al. 1999):

- Mechanical pre-treatment and primary sedimentation,
- Activated sludge process for removal of organic matter and sedimentation,
- Phosphate precipitation and final sedimentation.

Prior to 1995 the HIAS MWTP had no sludge digestion system. The sludge was untreated and landfilled (Fjaergard et al., 2006). Cambi® Technology was started in the HIAS MWTP in 1995. The installation was designed to treat 3600 dry MT per year. The raw sludge treated is a mixture of primary sludge (one third), secondary sludge (one third) and tertiary sludge (one third).

- ***HIAS MWTP Cambi Technology and Performance***

The goal of the Cambi® project in this MWTP was to reduce the quantity of sludge, produce more energy and operate a dryer (Fjaergard et al. 2006). From the final sedimentation tank, the sludge is pumped into a thickener. The thickened sludge is dewatered in a centrifuge to about 15-20% dry solids, and then pumped to the Cambi® installation. The latter includes one pulper, one hydrolysis reactor and one flash tank. The installation includes as well a steam boiler and a pressure control system.

After the thermal treatment, the hydrolyzed sludge is directly pumped into the anaerobic digester with a concentration of about 10-12% dry solids (Kepp et al., 1999). The anaerobic digester has a volume of 1500 m^3 and operates with a hydraulic retention time of 17 days (Kepp et al., 1999). All final residue generated at HIAS MWTP is used in agriculture (Kepp et al., 1999).

Table B-7 summarizes and compares the energy performance of the HIAS Cambi® installation with conventional anaerobic digestion. This data indicates that the Cambi® Technology produces about 1.5 times more biogas energy than conventional anaerobic digestion. The average VS destruction of a combined Cambi®-anaerobic digestion for this plant is estimated at 59% (Potts and Jolly, 2007).

It however not clear whether the comparison presented in Table 4-7 is based on the same energy recovery system (same efficiency) or not. According to this data, the HIAS plant has a net electricity production of 223 kW per year. About 533 kW of heat is also generated. The numbers presented here are equivalent to 0.07 kW of electricity and 0.17 kW of heat per dry MT of sludge treated. The conversion of those numbers into kWh gives approximately the following figures:

- 1.972 kWh per dry MT treated for the gross production,
- 0.3 kWh per dry MT treated for the electricity input,
- 1.672 kWh per MT treated for the net electricity produced,
- 4 kWh per dry MT treated for the heat produced.

Table B-7. Energy (kW) production from 3200 dry MT treated per year using the Cambi® Technology (Kepp et al. 1999).

Parameters	Cambi® Technology	Conventional Digestion
Biogas energy	1000	678
Electricity produced	263	210
Electricity to hydrolysis unit	24	/
Pumping sludge to digester	/	8
Final dewatering	4	8
Digester mixing (8 W/m^3) [a]	12	28
Net electricity	***223***	***175***
Recovered heat	***533***	***303***

[a] Based on gas mixing for the sake of comparison, HIAS uses propeller (1 W/m^3); "/" indicates not applicable

Cost Estimate

A report published by U.S. EPA (2006) indicates that capital costs and 2000-2001 O&M costs for Cambi® Technology installed at the HIAS MWTP was approximately US$6 million and US$396 per dry MT treated respectively. According to GVRD (2005) the installation cost for the Nigg Bay MWTP (15000 dry MT per year) in Scotland was estimated at US$9.5 million. It is however not clear what is included in those figures. For example, it is not clear whether or not those capital costs include both the technology cost and the installation cost. It is also not clear if the costs include the anaerobic digester.

Detailed cost estimates of the Cambi® Technology are not available in literature. Based on the information reported by U.S. EPA (2006) and GVRD (2005) the capital cost for the Cambi® Technology could be between US$650 and US$1700 per dry MT treated per year for an installation treating between 3600 and 15000 dry MT per year. This should be considered as a very rough estimate rather than an actual cost estimate.

B.3.1.1.2 BioThelys® Technology

Description

BioThelys® was developed by the French Company Veolia (Chauzy, 2003). Figure B.4 shows the flow diagram of the BioThelys process. This technology uses a single reactor to treat a feed with a solids concentration higher than 10% dry solids. The reactor operates at temperatures between 150°C and 180°C and pressures between 800 and 1000 kPa (8 and 10 bars). The retention time varies between 30 and 60 minutes. According to the vendor, this technology can reduce sludge production by up to 80%.

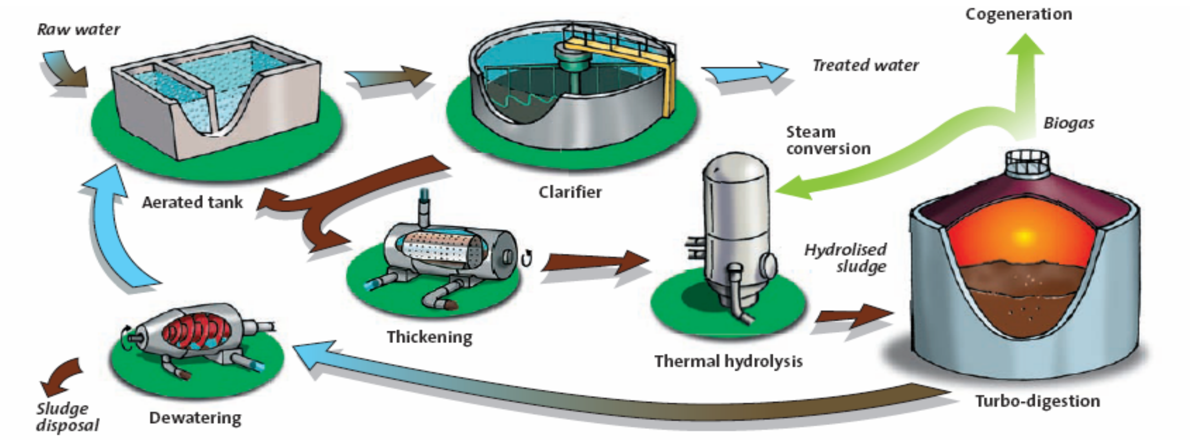

Figure B-4. Process flow diagram of a typical BioThelys® (From Veolia, 2007).

Examples of Full-scale Applications

The first functional full-scale BioThelys® installation has operated at Saumur in France since 1998 (Chauzy, 2006). This plant treats 1400 dry MT of sludge per year. Coupling BioThelysis with anaerobic digestion at the Saumur plant resulted in reducing the quantity (in kg/d) of wet residue for disposal by 44% (Chauzy et al., 2007).

A second installation operates at Château-Gontier, also in France, with a capacity of 1000 dry MT per year (Veolia, 2007). Detailed information about these two plants is not available. Attempts to obtain further information, such as cost and energy input/output, from the technology vendor were unsuccessful.

B.3.1.2 Cell Destruction

Cell destruction is a process that is used to destroy the cellular membrane of the microorganisms in the sludge, resulting in release of the soluble cell contents. The ruptured sludge solids are then treated in an anaerobic digester. In digestion, cell destruction requires more time than hydrolysis of the soluble cell contents, and is thus the rate limiting step. With cell destruction, the membranes are lysed more quickly than in digestion alone, the overall rate digestion is reduced. This results in more rapid and increased degradation of the volatile solids (VS) in the digester, reportedly up to 80-90%. Combination of a cell destruction process with anaerobic digestion produces more biogas than conventional anaerobic digestion alone. Examples of cell destruction technologies are MicroSludge^{TM}, ultrasonic treatment, ozonation, pulse electric fields and mechanical disintegration.

B.3.1.2.1 MicroSludge^{TM}

Description

This technology consists in mixing caustic soda (NaOH) and waste activated sludge containing 5-10% dry solids. The mixture is held for about one hour to weaken cell membranes. The process employs a high pressure homogenizer (about 800 kPa or 80 bars) or cell disrupter to provide an enormous and sudden pressure drop (from 800 to 40 kPa) to lyse the bacteria cells in the sludge. The liquefied sludge is then mixed with primary sludge and treated in an anaerobic digester with a detention time less than 9 days. Figure B-5 shows the simplified process flowsheet of MicroSludge^{TM}.

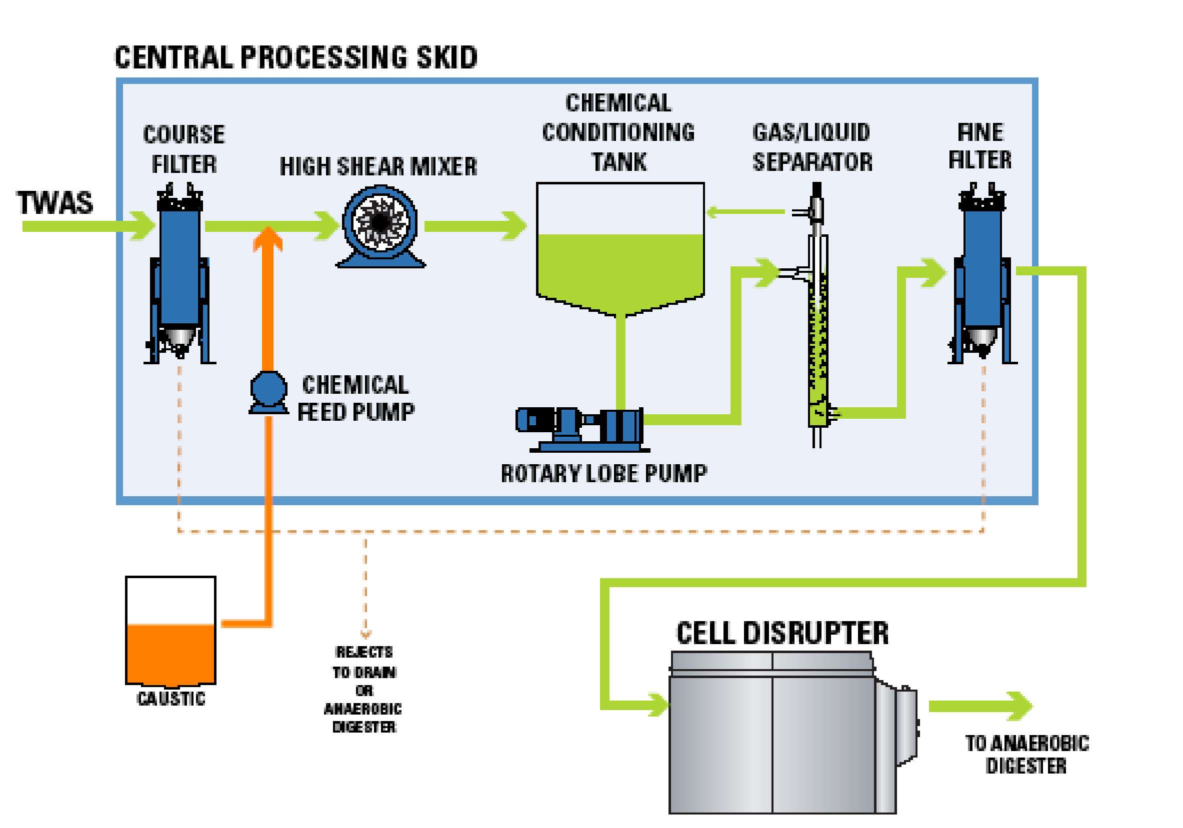

Figure B-5. Simplified process flow diagram of MicroSludge™ (From MicroSludge, 2007a).

According to the vendor (MicroSludge, 2007b), for each dry MT of waste activated sludge (WAS) at 80% VS, approximately 1358 kWh of electricity could potentially be produced by an engine at 30% biogas-to-electricity conversion efficiency. MicroSludge™ uses about 37% of the electricity produced, or 502 kWh per dry MT of WAS. This leads to a net electricity production of 856 kWh per dry MT of waste treated. These numbers are significantly higher than the electricity production reported for the Cambi™ process. The estimation is however based on an optimistic assumption, of 95%VS removal. Thus the actual quantity of energy that could be recovered will strongly depend on the actual %VS removed.

Examples of Full-scale Applications

The main vendor of MicroSludge™ process is Paradigm Environmental Technologies Inc. The first full-scale MicroSludge™ installation was demonstrated at the Chilliwack MWTP near Vancouver, BC, Canada in 2004. The second full-scale demonstration installation started at the Hyperion plant of Los Angeles County Sanitation District in October 2005. The technology is no longer in operation in those two plants. Testing was discontinued at the Hyperion plant when the energy balance was unsatisfactory due to difficulties in treating a number of different sludge types; the Chilliwack plant discontinued to re-evaluate the operation there following the Hyperion tests (Mavinic, 2007).

Cost Estimate

Cost estimate for MicroSludge^{TM} is summarized in Table B-8. This estimate was provided by U.S. EPA (2006). According to this source, the costs were obtained from the vendor. The capital cost provided here applies to a system processing approximately 189 m^3 (50000 US gallons) of thickened WAS per day. This cost does not include the fees for the installation of the equipment. The O&M cost estimate includes electricity, chemicals and maintenance. This cost is for processing a feed with concentration of 4 to 7% total solids (TS). The estimate assumes electricity is purchased at US$0.07 per kWh.

Table B-8. **Cost estimate for MicroSludge™ (After EPA, 2006).**

Technology	Capital cost (US$ million)	O&M cost (US$)
MicroSludge^{TM}	1.7 – 2.0	75 – 131 per dry MT of WAS [a]

[a] Cost converted from US short ton to MT, WAS indicates waste activated sludge

Assuming an average of concentration of 5.5% TS, the capital cost for processing daily 189 m^3 of WAS (about 10.395 MT per day) amounts to approximately US$488 per MT treated per year.

MicroSludge^{TM} looks very attractive due to the high potential for energy recovery. However, there are issues with process odors, equipment complexities and high organic/nutrient loads back to the headworks (Toffey, 2004). No odor problems exist when the feed is 100% WAS according to the vendor (MicroSludge, 2007b).

B.3.1.2.2 Ultrasonic Treatment

Description

This technology applies ultrasonic acoustic waves to wastewater solids to attain a very high pressure and temperature within the sludge. This results in the implosion of gas bubbles, which produces shear stresses that break up membrane walls of bacteria and other cellular matter (Hogan et al., 2004). Ultrasonic units are available in multiples of 1 kW, 2 kW, 4 kW, 8 kW, and 16 KW per probe. Ultrasonic treatment is typically applied to waste activated sludge (WAS) rather than primary sludge (PS).

The suitable feed concentration for optimal ultrasonic treatment has not been found in the literature consulted. However a feed with 10% dry solid was treated at the Mannheim Sewage Treatment Works (Bartholomew, 2002).

Examples of Full-scale Applications

The main vendor of Ultrasonic is EIMCO® Water Technologies. At least nine installations were built in different countries (Table B-9). Most of the installations are located in Europe and particularly in Germany. There have been also a number of full-scale trials including Avonmouth, Wessex, UK and Orange County Sanitation District (Hogan et al., 2004).

Table B-9. Full-scale Ultrasonic installations (Completed after Bartholomew, 2002).

WWTP Name	Location	Sludge	Capacity (P.E)	Ultrasound load (kW)	Start-up
Darmstadt MWTP	Germany	Mixed (PS-WAS)	180,000	16	2000
Süd Treatment W.	Germany	100%WAS	40,000	6	2000
Detmold	Germany	Mixed (PS-WAS)	95,000	14	2000
Mannheim MWTP	Germany	50%PS-50%WAS	725,000	24	2001
Rüsselsheim MWTP	Germany	Mixed (PS-WAS)	800,00	10	2001
Wiesbaden MWTP	Germany	35%PS-65%WAS	360,000	48	2002
Kävlingue MWTP	Sweden	25%PS:75%WAS	N.A.	N.A.	2002
Bad Bramstedt SW	Germany	N.A.	N.A.	N.A.	N.A.
Mangere MWTP	New Zealand	N.A.	80,000	432	N.A.

N.A. indicates information not available

Below are summarised the results observed in three of the installations (Bartholomew, 2002):

- In the *Süd Treatment Works,* the anaerobic digesters have experienced an average of 50% improvement in VS destruction. This resulted in 45% increase in biogas production.

- At the *Darmstadt WWTP,* improvements have been made to VS destruction from 44% to 55%. This resulted in an average increase in biogas of approximately 50% prior to treatment.

- At the *Mannheim Sewage Treatment Works*, the start-up of the ultrasonic pre-treatment, caused the VS destruction to increase to 70%. This resulted in an increase in biogas production of 45%. The additional biogas produced by ultrasonic treatment has resulted in an electricity generation of 1.2 MW. As a result, €285,000 (US$443,754) per year was saved by the plant. The drying capacity of the plant also dropped by 25%. These benefits have enabled plant to pay for the installation in 8 months.

Cost Estimate

The capital cost for ultrasonic treatment, based on a facility processing a 19,000-30,000 m^3/d (5-8 million gallon per day) of wastewater and treating 30% of the sludge daily is US$ 26,5000, while the O&M costs are between US$10,000 and 20,000 per year (U.S. EPA, 2006). The O&M costs assumptions include supervision, parts, and power. The power consumption of an ultrasonic unit is about 3.7 kWh per m^3 of wastewater processed (Nemw, 2001).

Based the above information and assuming that sludge production amounts to 0.08 dry solids kg per capita per day (Molton et al. 1986) and per capita wastewater production is equivalent to 0.3345 m^3/d (Monteith et al. 2006), an approximation of the costs and energy requirement for the ultrasonic process per MT of sludge treated was calculated. The results are summarized in Table B-10.

Table B-10. Energy input and cost estimates for the Ultrasonic process.

Technology	Value	Units
Sludge treated	498 - 786	Dry MT per year
Energy input	141	kWh per dry MT treated
Capital cost	533 - 842	US$ per dry MT treated per year
O&M costs	20 - 25	US$ per dry MT treated

B.3.2 Sludge-to-Syngas Processes

Sewage sludge can be converted in energy-rich gases that can then be used to generate electricity. The technologies developed to perform such treatment are based on gasification, incineration and pyrolysis processes.

B.3.2.1 Gasification

Gasification is a process that takes place in two steps. In the first step, the volatile fraction of the solids is transformed, in absence of air, into a carbon-rich substance called "char". This transformation occurs at a temperature around 600°C or less. In the second step, the char is gasified in the presence of oxygen or air. This reaction produces a gas called syngas. The latter typically contains nitrogen (N_2), carbon dioxide (CO_2) and carbon monoxide (CO) and hydrogen (H_2). Syngas can be used as fuel to generate electricity and heat.

Examples of gasification technologies are the KOPF Gasification Technology and the EBARA fluidized bed gasification technology.

Kopf Gasification Technology

Description

The main components of the KOPF gasification technology are: a solar drying unit, a fluidized-bed gasification unit, a gas engine unit for energy recovery and a post combustion chamber (KOPF, 2007a). Figure B-6 shows the detailed process flow diagram of the KOPF gasification technology (KOPF, 2007b).

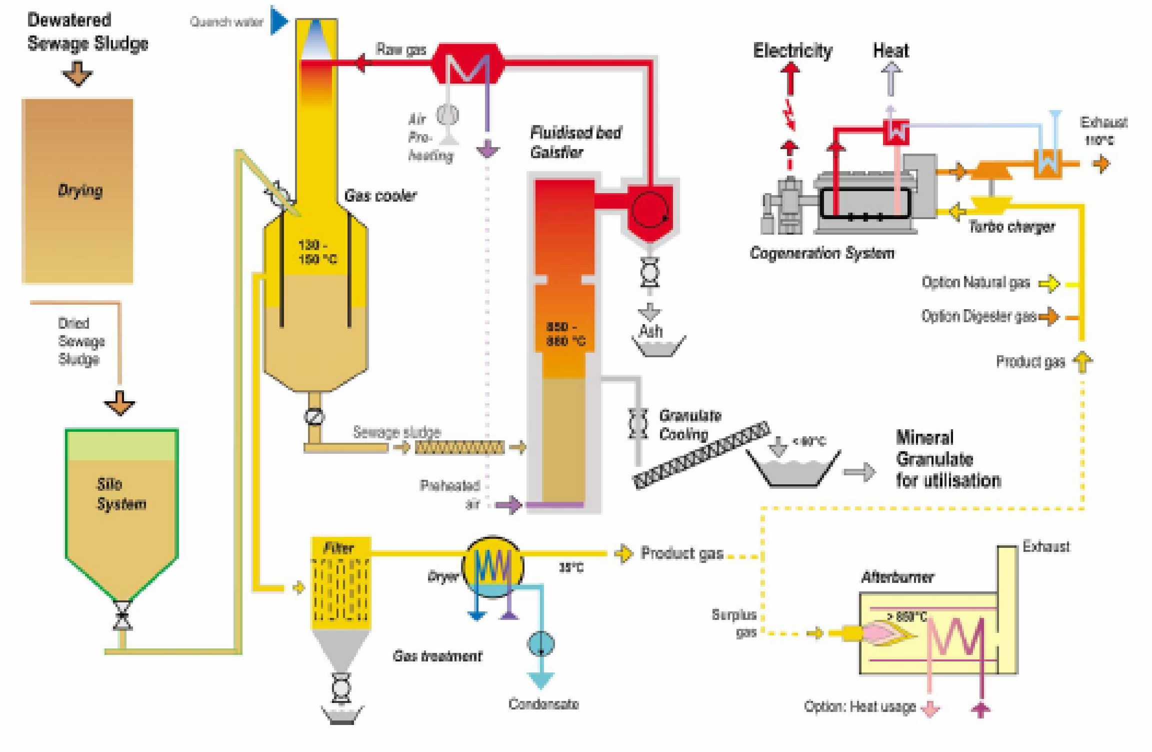

Figure B-6. Process flow diagram of the KOPF Gasification Technology (From KOPF, 2007b).

Solar drying unit
This unit is used to dry the wet digested sludge to a solid content of between 70 and 85%. The residence time of the solids in the unit is between 2 to 8 weeks depending on the weather conditions. The electricity required for operating the unit is supplied by a photoelectric system. The dried solids are transferred by conveyor belt to the fluidized-bed reactor.

Solar drying systems are being used more commonly around the world because they offer a great potential for energy saving potentiality. Currently there are 36 Thermo System ™ solar dryers in Europe, 3 in the U.S., 2 in Brazil and 2 in Australia. Examples of installations are Keowee Key in South Carolina (Mooney et al. 2003; Bowen 2003), the City of Rogue River in Oregon (The Dryer Partnership Inc. 2007), Renquishausen in Germany, Houdan in France, Krems in Austria, Tiers in Italy, Fuessen in Germany and Bad Vöslan in Austria (Thermo-System 2007). With the large number of installations in Europe it seems that solar drying can be used in European climate. The time required for drying will be longer in winter than in summer.

Fluidized-bed gasification unit
This unit is the key component of the KOPF gasification technology. The reactor is a stationary fluidized bed operating at a temperature up to 900^{o}C. The reactor has a retention time of 30 minutes (KOPF, 2007a). Pre-heated air to around 400^{o}C is used to ensure the fluidization of the bed. Inside the reactor, dried solids are converted into inert granules and combustible gas. The gas is recovered and cooled to a temperature below

150°C and dried. The cold gas efficiency is between 65 to 70%, depending on the degree of drying. For starting up the plant, natural gas is required. After the start-up phase, no external fuel is needed.

Gas engine unit
The gas engine unit produces about 70 kW of electricity. 15 kW is used to operate the gasification installation and the remaining 55 kW is used to cover the energy demand of the sewage works. About 140 kW of thermal energy is also recovered. This energy is used to heat the digester towers.

Post combustion chamber
The post combustion chamber is used to dispose of the surplus of gas that can not be utilised. The gas is combusted at a temperature of 850°C.

Examples of Full-scale Applications – Balingen SW, Germany

Most of the full-scale gasification installations are operating in Europe, and mainly in Germany. Balingen Sewage Works is one of the plants operating in Germany (Kopf, 2007). This plant has been in operation since August 2004 and uses the Kopf Gasification Technology (Kopf, 2007).

○ *Balingen SW Capacity*

The Balingen Sewage Works treats an annual wastewater flow of 10 million m^3. The plant is designed for a connecting capacity of 125,000 P.E. In order to utilize the energy content of the digested sludge, the local association for wastewater cleaning ordered a sludge gasification plant. In August 2004, a fluidized-bed gasification plant, manufactured by the German company KOPF was constructed in the plant for processing the digested biosolids.

○ *Balingen KOPF Gasification Technology and Performance*

The Balingen plant processes about 230 kg of dried sewage sludge granulate per hour. Depending on the degree of drying, this is the equivalent of 160 to 180 kg dry substance of sewage sludge. The mineral granule produced amounts to 85 kg per hour. The plant produces about 300 m^3 of exhaust per hours. The material and energy balances of the gasification installation are summarized in Table B-11. Based on this data, 0.5 kWh of electricity is produced per kg TS treated. Only 0.1 kWh per kg TS treated is used for the gasification installation and the remaining 0.4 kWh is used to cover the sewage plant demand. This energy balance looks quite attractive knowing that energy could also be recovered from the biogas in case the sludge is initially pre-treated in an anaerobic digester.

Table B-11. Material and energy balances of the KOPF fluidized-bed gasification Process (KOPF, 2007b).

Mass Balance Parameters [a]	Mass	Energy Balance Parameters [b]	Energy
Dewater digested sludge (32% DS)	1000 kg	Dried digested sludge	8.3 MJ/kg TS
Dried sludge (80% DS)	400 kg	Gas production	1.6 m^3/kg TS
Water vapor	600 kg	Heat from gasification	0.1 kW/kg TS
Granulate	160 kg	Electricity produced (CHP)	0.5 kWh/kg TS
Gas	510 m^3	Heat produced (CHP)	0.9 kWh/kg TS

[a] Assume a gasification with 290 m^3 of ambient air; [b] Assume cold gas efficiency is 70%, CHP efficiency is 30% for electricity and 60% for heat, lower heating value is 1 kWh/m^3.

EBARA Fluidized Bed Gasification Technology

The EBARA TwinRec process operates with a temperature between 500 and 800°C (Arena and Mastellone, 2005). Figure B-7 shows the process diagram of the Twin Rec. An example of plant is located in Aomory, Japan, with a capacity of 20 MT per hour. This plant treats 30% sewage sludge and 70% shredder residues (Steiner et al. 2002).

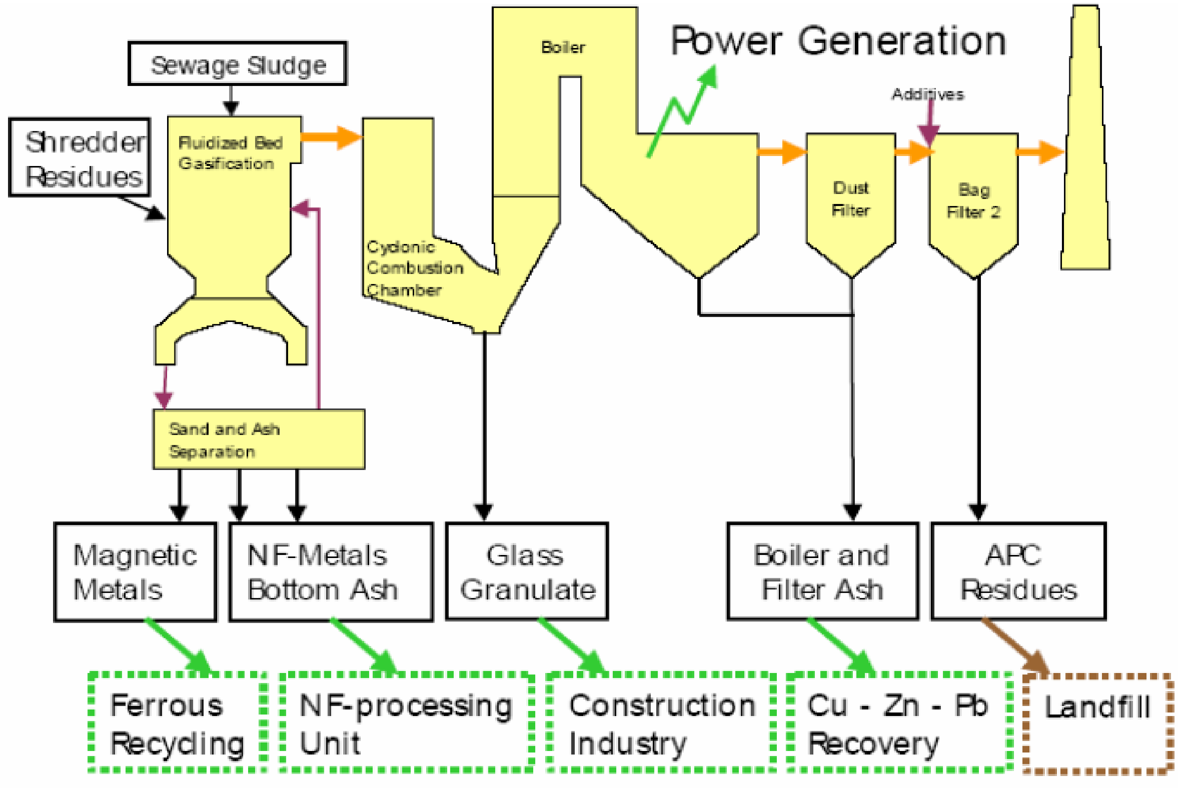

Figure B-7. Schematic representation of the Twin Rec Process (From Arena & Mastellone, 2005).

According the Steiner et al. (2002), as of spring 2002 six commercial TwinRec units were in operation. These commercial plants are designed to treat shredding residues, sewage sludge, fly ash, waste plastics, liquid waste, medical waste and municipal solid waste. There is very limited information about the full-scale plants using the EBARA' technology.

B.3.2.2 Incineration

Incineration, also called combustion or thermal oxidation, is the process of combusting organic waste at high temperature in the presence of oxygen (air). This process involves five main steps: 1) Dewatering, 2) Drying, 3) Combustion, 4) Air pollution control equipment, and 5) Ash management.

Dewatering is used to increase the solids concentration of the feed to between 15 and 35%. Drying raises the sludge temperature to the point that water in the solids evaporates. Combustion is used to destroy the volatile fraction in the sludge. This is an exothermic reaction that results in the production of inert ash and hot gases. Energy can be recovered from the hot gases.

The ash produced during the combustion can be landfilled or it can be used as additive for cement and brick manufacturing. Ash may also be used for phosphorus recovery.

Two types of incineration technologies are used for sludge combustion: older technology multiple hearth furnaces (MHFs) and newer technology fluidized bed furnaces (FBFs).

Incineration is a commercially proven technology. The U.S. EPA (2003) reported that in 1993, 343 biosolids incinerators were in operation in the United States. Of these, approximately 80% were MHFs and 20% were FBFs. Use of FBFs technology for wastewater solids incineration has considerably increased the last 10 years because they are more efficient, more stable and easier to operate than MHFs (U.S. EPA, 2003). Several plants in North America have replaced their MHF installations by a FBF installation (Dangtran et al., 2002).

Thermylis® HTFB

Description

Thermylis® HTFB is a High Temperature Fluidized Bed Incineration Technology developed and commercialized by Degrémont (Degrémont, 2007). This process operates with a temperature around 840°C. Figure B-8 shows the flowsheet of Thermylis®.

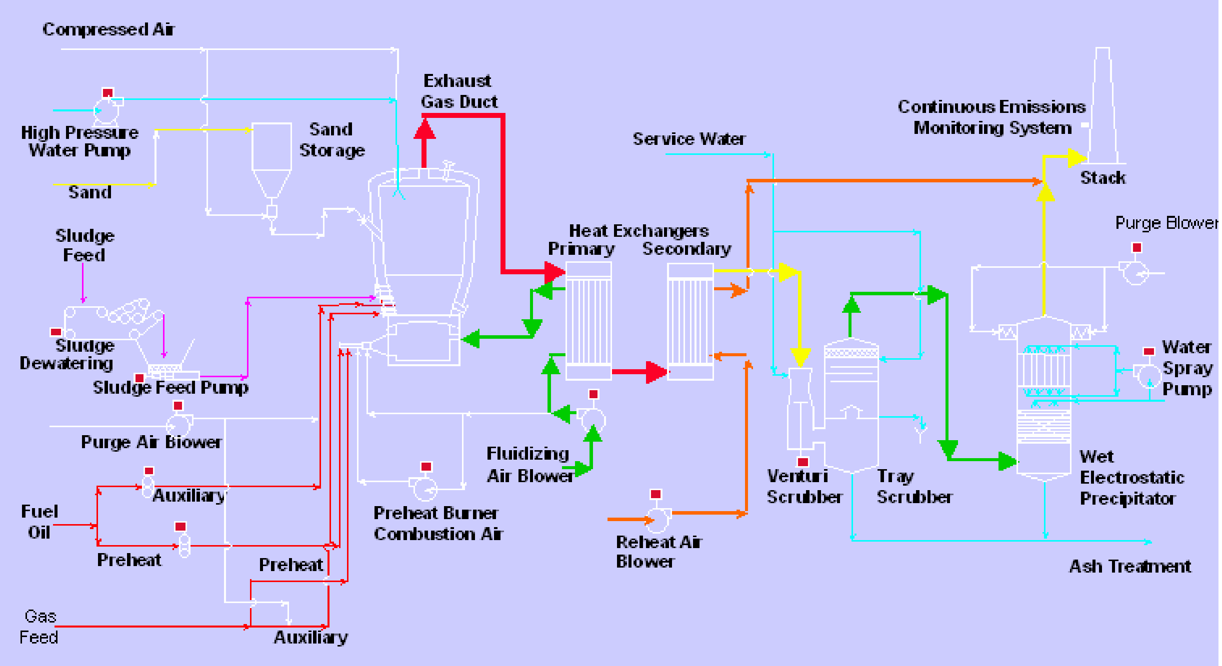

Figure B-8. Flowsheet of Thermylis® (From Degrémont, 2007).

Examples of Full-scale Applications

About 14 installations have been constructed from 1995 to 2007 in North America and France. The name and location of the plants are summarized in Table B-12. Three new installations are planed in Lakeview, Ontario, Canada in 2008 with a capacity of 110 dry MT per day each.

Table B-12. Full-scale Thermylis® installations (From Degrémont, 2007).

	Location	Capacity (dry MT per day)	Start-up
Pfizer	CT-US	18	1995
Bayshore	NJ-US	27	1995
Morton International	MS-US	76	1996
Greensboro	NC-US	60	1996
Camden County	NJ-US	10	1996
Northwest Bergen	NJ-US	27	1999
Little Miami	OH-US	72	2000
District d'Elbeuf	France	11	2003
Valence	France	14	2003
Puerto Nuovo	PR-US	64	2004
Ypsilanti	MI-US	76	2005
Lakeview I	ON-Canada	110	2006
Cobb County I and II	GA-US	53 x 2	2007

Other Variants – Knostrop STW, UK

Hand-Smith (1999) reported that several incineration plants were operating in the UK in 1999 and others were in their commissioning phase. Thames Southern Water (TSW) and Knostrop are some the installations in operation.

- **Knostrop STW Capacity**

Knostrop STW treats sewage from the Leeds conurbation. It has a capacity 940,000 P.E.

- **Knostrop STW Incineration Process and Performance**

The incinerator burns 3.3 MT of dry solids per hour at a temperature higher than 850^{o}C. The process used for sludge processing at the Knostrop STW includes;

- A dewatering unit to produce a sludge cake.

- Sand FBF furnace for combusting the sludge cake. The bed is fluidized by hot air, which evaporates the remaining water and incinerates the sludge to an inert ash.

- A heat recovery systems used to recover heat from the flue gases. The heat is used to pre-heat the combustion air and to generate steam. The steam is then used to pre-dry the feed sludge and generate electricity in a steam turbine.

- A series of sophisticated processes including an adsorption stage for the removal of mercury from the flue gases.

- **Cost estimate of the Knostrop STW Incineration Process**

The cost of the incineration project was 32 million British Pounds (Hand-Smith, 1999). Assuming 1 GBP is equal to US$2, the capital cost is then approximately equivalent to US$64 million. Based on the plant capacity the capital cost can be estimated at approximately US$2,214 per dry MT of sludge treated per year.

B.3.3 Sludge-to-Oil Processes

Sewage sludge can be converted to oil, which then can be used as an industrial fuel in boilers to produce steam, or it can be combusted in an engine to produce electricity (Bridle, 2004). The oil can also be used for biodiesel production (Salter, 2006; Dufreche et al., 2007; Liang et al., 2007). The processes that can be used to produce oil from sludge include pyrolysis and hydrothermal treatments.

B.3.3.1 Pyrolysis

Pyrolysis is the degradation of waste in the absence of air. During pyrolysis materials are converted to char and syngas. The syngas may be converted to oil depending on the operating conditions. There are two categories of pyrolysis; slow pyrolysis and fast pyrolysis. Slow pyrolysis does not produce oil. It takes place at temperatures below 315°C. This process operates with a long vapor residence time in the reaction zone.

Fast pyrolysis takes place between 425 and 538°C. This process has a shorter vapor residence time and produces oil. The main features of fast pyrolysis are: 1) drying the sludge to less than 10% water in order to minimize the quantity of water in the final oil stream, 2) finely grinding of sludge to around 2 mm in the case of fluidized bed reactors, 3) non-pressurized reactor, 4) controlled reaction temperature, 5) vapor residence time less than 2 seconds, 6) rapid cooling and condensation of the vapor to produce bio-oil, and 7) collection and separation of solid by-product.

Commercial pyrolysis technologies for sewage sludge treatment are EnerSludge™ Technology (Bridle, 2004; Bridle and Skrypski-Mantele, 2004) and SlurryCarb™ Technology (EnerTech, 2006).

EnerSludge™ Technology

Description

EnerSludge™ has been developed over the past 15 years, with extensive pilot plant demonstration programs conducted in Australia and Canada (Bridle et al. 1989; Gough et al. 1991). This technology was developed by Environmental Solutions International (ESI).

Examples of Full-scale Applications – Subiaco MWTP, Australia

The first commercial demonstration installation was constructed at the Subiaco MWTP in Perth, Western of Australia. This installation is described below.

- ***Subiaco MWTP Capacity***

The Subiaco Sewage Works treats domestic wastewater mainly collected from the Perth central metropolitan area. The plant is designed to treat up to 61,400 m^3 per day, equivalent to a population of 350,000 people. The wastewater is predominantly from household kitchens, bathrooms, toilets and laundries.
The works includes preliminary treatment, primary sedimentation, conventional activated sludge with nutrient removal, final sedimentation and sludge treatment (combined primary sludge and thickened WAS).

- ***Subiaco MWTP EnerSludge™ Performance***

EnerSludge ™ was operated at the Subiaco MWTP from September 2000 to December 2001. Figure 4-9 depicts the process flow diagram of the installation.

After dewatering and drying to 95% TS, the pellets are fed to the pyrolysis unit. Inside the unit, the pellets are heated to 450°C in the absence of oxygen at a low pressure (0,015 bar) producing char and syngas. The char and syngas are re-contacted in a second reactor, also operated at 450°C to facilitate the catalyzed vapor phase reactions that refine the syngas, producing mainly hydrocarbons.

The refined vapors are finally condensed to 50°C in a direct contact spray condenser using cooled reaction water (RW) as cooling medium. The condensed products are separated in an oil-water separator, with the oil stored prior to use as a fuel off-site. The char is cooled down to 70°C and then used together with RW and non-condensed gas (NCG) as fuel in a Hot Gas Generator (HGG). The HGG unit, a fluidized bed combustor, produces heat and flue gases. The heat produced is used on-site for sludge drying.

The flue gas, at a nominal temperature of 850°C transfers heat to a closed-loop of drying air in an air-to-air heat exchanger (not shown in Figure B-9). Cooled flue gas from the heat exchanger is then cleaned in a venturi scrubber and SO_2 scrubber, before discharging to the atmosphere, via the plant stack. The final residue (ash) can be used for brick production (Ashford, 2007).

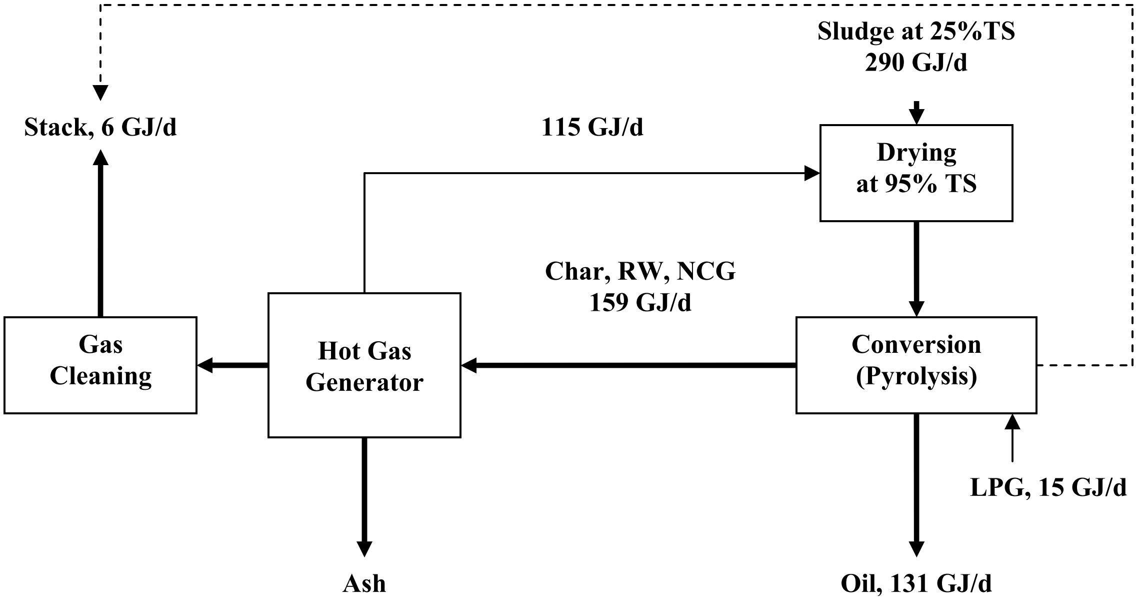

Figure B-9. Simplified process flow diagram of the EnerSludge™ Technology experimented at the Subiaco WWTP (Adapted after Bridle 2004).

When processing 15 dry MT per day of sludge, the facility requires 15 GJ of liquefied petroleum gas (LPG) per day or 1 GJ of LPG per dry MT. The plant produces a net energy of 7.7 GJ per MT of dry sludge treated in the form of oil (Bridle, 2004). Overall 45% of the energy in the biosolids is converted to oil. According to Bridle (2004), if the oil were combusted in a diesel engine to produce electricity, the process would generate 925 kWh per dry MT of sludge processed. This assumes a conversion efficiency of about 38%. From the above information it appears that quantity of electricity that could be recovered with EnerSludge™ is about 2.3 times the net quantity of the electricity generated from gasification using the KOPF technology, a process operating with a higher temperature (900°C). In addition to the oil, about 7.67 GJ of heat per dry MT treated can be recovered from the hot gas.

- ***Cost estimate of the of the Subiaco MWTP Enersludge^{TM} Process***

Detailed cost estimate for Enersludge^{TM} was reported by Ashford (2007), the General Manager of ESI. Based on the information provided by Ashford (2007) EnerSludge^{TM} ;

- Capital cost could be roughly between U$1640 and U$1954 per dry MT treated per year for an installation treating 5475 to 9125 dry MT per year.

- O&M costs could be roughly between US$86 and US$113 per dry MT treated for an installation treating 5475 to 9125 dry MT per year.

SlurryCarb^{TM} Technology

Description

SlurryCarb^{TM} was developed and patented by Atlanta-based EnerTech Environmental. Basically this technology works in seven steps including, sludge preparation, slurry pressurization, slurry heating at a temperature between 270°C and 325°C, reaction, dewatering/drying, filtrate processing and recycle and fuel utilization (Figure B-10).

SlurryCarb^{TM} converts sludge into a fuel called E-fuel and CO_2 gas. The E-fuel is the carbonized product of the sludge. The product can be recovered either as slurried E-fuel or as Dried E-fuel. These two materials can serve as fuel in cement kilns (EnerTech, 2006).

Application

EnerTech Environmental built a pilot plant in the city of Ube in Japan. This plant operated from 1997 to 2000 as demonstration project. A new full-scale installation is now being constructed at the Rialto MWTP in California. Currently, the Rialto wastewater treatment produces about 27 MT of sludge per day.

The start-up of the SlurryCarb^{TM} installation at the Rialto WWTP is anticipated early in 2008. The facility has been designed to treat 687 MT per day of sludge, containing 10 to 31% dry solids. Based on the design capacity, it is anticipated that the plant will recover approximately 120 MT of dried pellets per day that can be used as renewable fuel. The energy value of the product is expected to be roughly 15.6 MJ per kg (Govt-NZ, 2006). The plan for the Rialto project is to market the pellets to a near by cement plant.

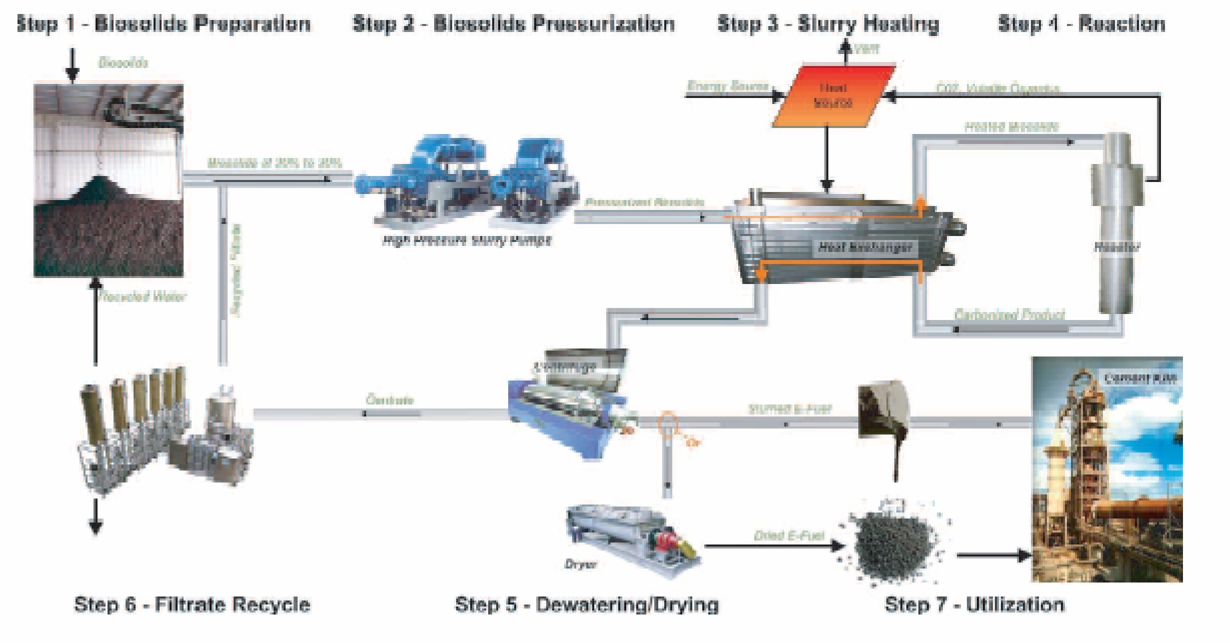

Figure B-10. Flow diagram of the SlurryCarb™ process (From Bolin et al. 2007).

Cost Estimate

The capital cost of the Rialto installation is expected between US$60 – 100 million (Govt-NZ, 2006). The operating cost could be high as well because of the process energy requirement and labour cost. As reported by Govt-NZ (2006), the installation could need 38 million of BTU per hour (use of natural gas) and 4 MW of electricity. The plant will hire 20 to 25 staff, each earning US$20 per hour.

In order to compare the SlurryCarb™ process with other technologies, a rough estimation of energy input, energy output and cost was performed using the information above. The results are summarised in Table B.13. This estimation assumes the following conditions:

- 250,755 MT of sludge per year with 22.5% dry solids,
- 7.5 hours of work per day, 5 days of work per week, 52 weeks per year,
- US$0.08 per kWh

The results in Table B-13 shows that the energy output (energy contain of the pellets) is higher than the total input. Overall, the process will produce 6% more energy than it will consume. A different estimation performed by Bolin et al. (2007) concluded that the SlurryCarb™ process at Rialto will produced 44% more energy than it will consume.

If equipment maintenance costs were considered, the O&M costs would be higher than the operating costs indicated in Table B-13. The O&M costs would probably be around US$100 per MT.

Table B-13. Energy and costs estimates for the Rialto SlurryCarb™ project.

Parameter	Value	Unit
Dry solid treated [a]	56420	Dry MT per year
Electrical energy input	621	kWh per dry MT
Heat energy input	91	kWh per dry MT
Total energy input	712	kWh per dry MT
Raw energy output [b]	758	kWh per dry MT
Capital cost	1063 – 1772	US$ per dry MT per year
Operating cost-Labour	13.82 – 17.28	US$ per dry MT
Operating cost-Energy	57	US$ per dry MT
Operating cost-Total	70.82 – 74.28	US$ per dry MT

[a] Design capacity, [b] energy content of the pellets

B.3.4 Sludge-to-Liquid Processes

B.3.4.1 Super Critical Water Oxidation

Super critical water oxidation (SCWO) is a process that occurs in the water phase at temperatures and pressures above the super critical point of water. The process has been under development for more than twenty years (Svanström et al., 2004). It can be used to oxidize digested as well as undigested sludge. It operates at a temperature around 374°C, a pressure higher than 22,000 kPa (220 bars) and at low dry matter concentration, e.g. 3% (Levlin, 2004). SCWO process requires pure oxygen. Pure oxygen is usually produced using cryogenic air separation (Bernstein, 1999), which requires energy for operation.

This process is capable of removing more than 99.99% of the chemical oxygen demand (COD) present in the sludge within 1 minute (Stendahl and Jäfverström, 2004). The COD is converted to carbon dioxide. The effluent from the SCWO reactor is a slurry of inorganic ash in a water phase. Inorganic components like phosphates and/or coagulants can be recovered from the effluent (Stendahl and Jäfverström 2004). Table B-14 shows the composition of SCWO residue measured in four Swedish MWTPs by Stark (2005). The ash P_2O_5 content varies between 6.3 and 18.4 %. This indicates that it is indeed possible to recover P from the SCWO slurry.

Table B-14. Composition of SCWO residue from 4 Swedish MWTPs (Stark, 2005).

Parameters	Bromma	Borlänge	Karlskoga1	Karlskoga2
Dry mass (%)	5.4	0.7	3.7	3.4
Al_2O_3 (% of dry solids)	11.7	3.6	16.3	4.5
CaO (% of dry solids)	6.2	6.0	4.0	4.3
Fe_2O_3 (% of dry solids)	23.3	12.9	25.0	19.3
K_2O (% of dry solids)	0.48	0.94	0.45	0.61
MgO (% of dry solids)	1.0	1.1	0.8	0.92
P_2O_5 (% of dry solids)	10.9	8.2	18.4	6.3

During sludge treatment with SCWO, energy can be recovered directly by heat exchange in the reactor, or from the effluent leaving the reactor. The heat in the effluent is transferred to a water stream through a heat exchanger unit. The way energy is recovered with SCWO is different from that used during incineration. As indicated in section IV.2.2.2, during incineration heat is recovered from the hot flue gas.

Advantages of SCWO compared to incineration are: 1) no need to dewater the sludge at very high dry matter concentration and 2) low operating temperature. The main drawback of SCWO is that only heat can be recovered. This may be a significant limitation of the process because wastewater treatment plants need both electricity and heat to operate, but mainly electricity.

Examples of SCWO technologies are Aqua Reci® Technology (Stendahl and Jäfverström, 2004), Aqua Citrox® Technology (Gidner and Stenmark, 2001) and the Athos® Technology (Veolia, 2007). Only the last technology will be described in this section. The other two will be described later in Chapter 5.0 because there are classified as emerging technologies.

Athos® Technology

Description

The Athos® Technology operates at a temperature around 250°C, a pressure of 5,000 kPa (50 bars) and requires pure oxygen (Guibelin, 2004). Figure B-11 shows the simplified process diagram of the Athos® Technology.

Energy is recovered via the effluent from the reactor. The heat in the effluent is transferred to a water stream through a heat exchanger unit as shown in Figure B-11.

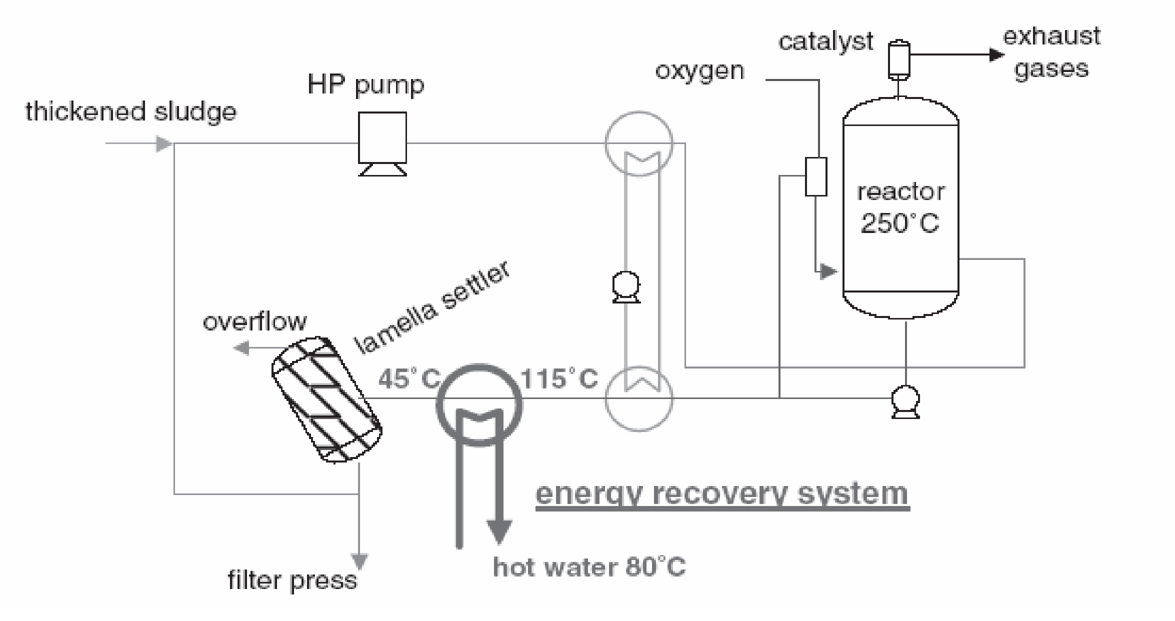

Figure B-11. Simplified process diagram of the Athos® Technology (From Guibelin, 2004).

The energy requirement of the process operation is 500 to 800 kWh per dry MT and 800 kg O_2 per dry MT (Guibelin, 2004). Assuming the production of 2 tonnes of oxygen requires 1 MWh of electricity, then 0.4 MWh per dry MT will be needed to produce 800 kg of oxygen per dry MT. The total electricity requirement is therefore between 900 kWh and 1,200 kWh per dry MT. The estimated energy recovered is about 1,680 kWh per dry MT. This indicates that the process could produce excess energy.

Application

The vendor's website indicates that up to 5 full-scale reference plants were constructed from 1998 to 2004 in Europe (Veolia, 2007). The plants are located in Toulouse-Gisnestous in France (50,000 P.E), Brussels-North in Belgium (1.1 million P.E), Trucazzano in Italy (300,000 P.E), Epernay in France (150,000 P.E) and Seine-Aval in France (15,000 P.E). Detailed operating or financial information regarding those plants is not available.

Appendix C:

Detailed Descriptions of Emerging Energy and Resource Recovery Technologies

C.1 RESOURCE RECOVERY

C.1.1 Phosphorus Recovery

C.1.1.1 KREPO Technology

Description

KREPO is an acronym for Kemwater REcycling PROcess (Hultman, 1999; Hultman, et al. 2001). This technology developed in Sweden uses heat, pressure and sulfuric acid to dissolve phosphates, metals and a large fraction of organic compounds from thickened sludge.

Figure C-1 shows the flow diagram of the KREPO Technology. The main steps involved in this technology are; acidification, heating, hydrolysis, organic sludge separation and precipitation. The raw sludge is initially thickened to 5% dry solids in a centrifuge.

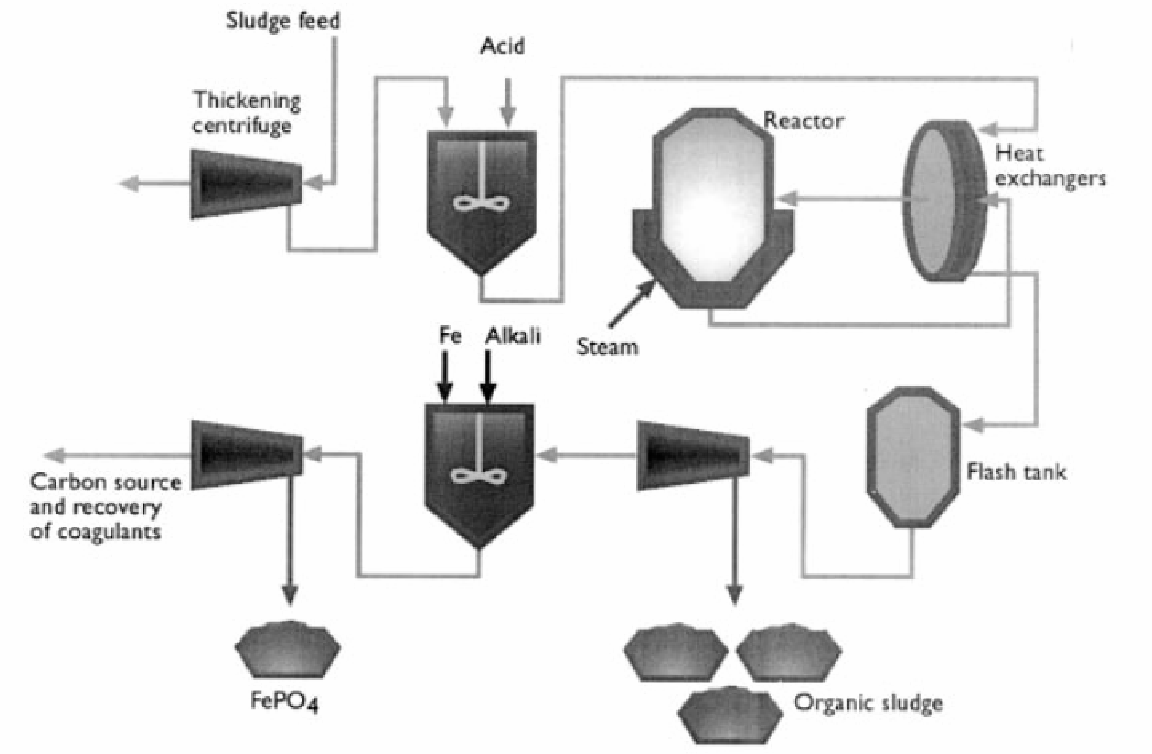

Figure C-1. **Simplified Flow Diagram of the KREPO Technology (From Hansen et al., 2000).**

Thickened sludge is mixed with H_2SO_4 in order to keep the pH below 2. This dissolves almost completely the heavy metals and phosphates. The acidified sludge is heated at a temperature between 100 and 110°C and a pressure of 3.6 bars in the reactor. The organic sludge is separated from the soluble phase which contains about 90% of the original amount of P. phosphorus in the soluble phase is precipitated as iron phosphate ($FePO_4$). Precipitation is achieved by adding to the soluble phase NaOH, to increase the pH to around 3, and ferric salt as precipitant. A final centrifugation is applied to recover the precipitated phosphorus.

Application

A full scale KREPO Technology system was operated in 1995 at the Öresundsverket MWTP in Helsingborg Sweden (Hansen et al. 2000). Table C-1 summarizes the resources used and produced by the KREPO system at the Öresundsverket STP.

Following the experience of the Öresundsverket MWTP, the city of Malmö planned a full-scale KREPO plant (SCOPE, 2001), although the plan was later abandoned (Hultman et al., 2003).

Table C-1. Resources used and produced from sludge treatment at the Öresundsverket MWTP with the KREPO Technology (Hansen et al., 2000).

Resources used	Quantity (per dry MT)	Resources produced	Quantity (per dry MT)
Sulfuric acid	200 kg	Phosphorus	27 kg
Sodium hydroxide	200 kg	Ferric phosphate/hydroxide	515 kg
Magnesium hydroxide	12 kg	Ferrous iron	38 kg
Ferric chloride	390 kg		
Polymers	2 kg		
Energy	463 kWh		

C.1.1.2 Kemicond^{TM} Technology

Description

Kemicond is an acronym for Kemira sludge conditioning. This technology was developed in 2003 by modifying the KREPO Technology (Berg and Shaum, 2005) described earlier. The process is used to treat raw sewage sludge. Figure C-2 shows the flow diagram of Kemicond^{TM} Technology. It consists of a chemical treatment by H_2SO_4 and hydrogen peroxide (H_2O_2), followed by a two stage dewatering unit (Cornel et al. 2005). The contact time between the chemical and the sludge is 40 to 60 minutes. During the chemical treatment, metals such as iron phosphate and hydroxides are dissolved. The peroxide oxidizes the dissolved iron (II) into iron (III). The dissolved phosphate can then be re-precipitated as ferric phosphate.

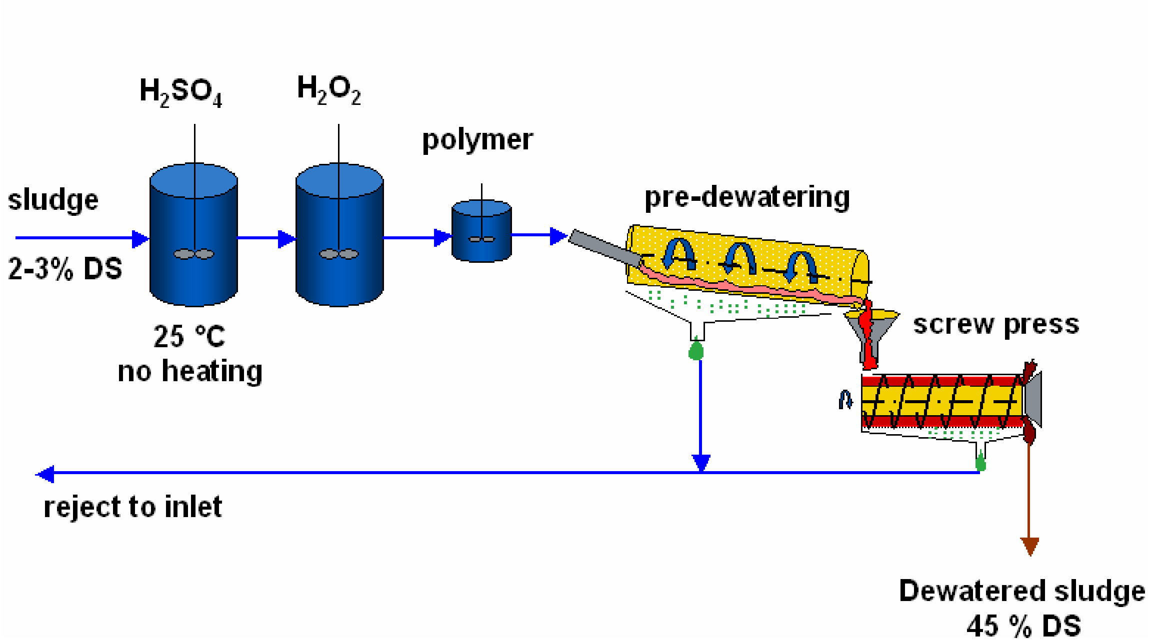

Figure C-2. Flow diagram of the Kemicond™ Technology (From Karlsson, 2007).

Application

The first plant that used the Kemicond^{TM} technology started up in Stockholm, at the Käppala MWTP. This plant treats wastewater of approximately 500,000 P.E. Because the main purpose of the Käppala project was to improve sludge dewatering and reduce sludge transport (Manhem and Palmgrem, 2004), phosphorus recovery was not investigated in detail. In the period from June 2003 to till March 2004, Kemira conducted several pilot scale tests in the plant. The goal of the tests was to determine the influence of chemical doses on the process performance. A detailed report of the investigations has been published by Nikolic and Karlsson (2005). The estimated chemical consumption for the Käppala MWTP is reported in Table C-2. The energy required to treat the 8,000 dry MT of solids is about 1,500 MWh per year (Manhem and Palmgrem, 2004). This is equivalent to 187.5 kWh per dry MT of solids treated.

Table C-2.Chemical consumption of the Kemicond™ technology (Manhem & Palmgren, 2004).

Chemical	Amount (kg per MT of sludge)
Sulphuric acid (H_2SO_4)	150-300
Hydrogen peroxide (H_2O_2)	25-60
Sodium hydroxide (NaOH)	7-40
Polymer	4-7

In June 2006, Oulun Vesi Water Utility in Oulu Finland, chose Kemicond for handling sludge from the city's waste water treatment processes. The technology will be installed at the Taskila MWTP which has a capacity of 150,000 P.E (Kemira 2006a).

Cost Estimate

A cost estimate of the Kemicond^{TM} technology for treating 8,000 dry MT of solids per year was reported by Kemira (2007b) the technology vendor. The investment for storage tanks, reactors, dosing equipments and process control are estimated at about €1 million (about US$1.5 million). The cost of chemical needed varies between €35 and €65 per dry MT of sludge treated (about US$53 to $98 per dry MT). The estimated investment cost per dry MT of sludge treated is approximately US$186.

C.1.1.3 Seaborne Technology

Description

The Seaborne technology was developed in Germany by the Seaborne Environmental Research Laboratory (Berg and Shaum, 2005). This technology is based on a combination of several processes, including incineration, acid treatment, desulphurization, methane production, heavy metals separation, and struvite precipitation

The Seaborne technology can basically be described in three steps. In the first step, metals and nutrient are dissolved by lowering the pH with an acid solution. The organic residual and the soluble compounds are separated using a centrifuge. The organic residual is incinerated. In the second step, hydrogen sulfide in digester gas is used to precipitate the metals from the centrate. This allows purifying the digester gas. The metals are separated by filtration. In the third and final step, sodium hydroxide and magnesium oxide are added to the filtrate. This results in the precipitation of struvite (MAP).

The main advantages of this technology are 1) the recovery of multiple nutrients with apparently no heavy metals and organic pollutants and 2) H_2S-free biogas (Müller et al., 2007). The complex technology however requires more unit processes than other technologies.

Application

The first large scale Seaborne pilot plant was built between 2005 and 2006 at the Gifhorn wastewater treatment in lower Saxony, Germany. The Gifhorn plant has a capacity of approximately 50,000 P.E. The Seaborne installation has a capacity of 1000 dry MT of solids per year (Müller et al., 2007). Figure C-3 shows the process flowsheet of the pilot plant. A fourth step is used to recover surplus nitrogen as ammonium sulfate in a stripping tower.

This demonstration plant has provided a of lot experience that will be helpful for future commercial installations. The process evaluation is still ongoing.

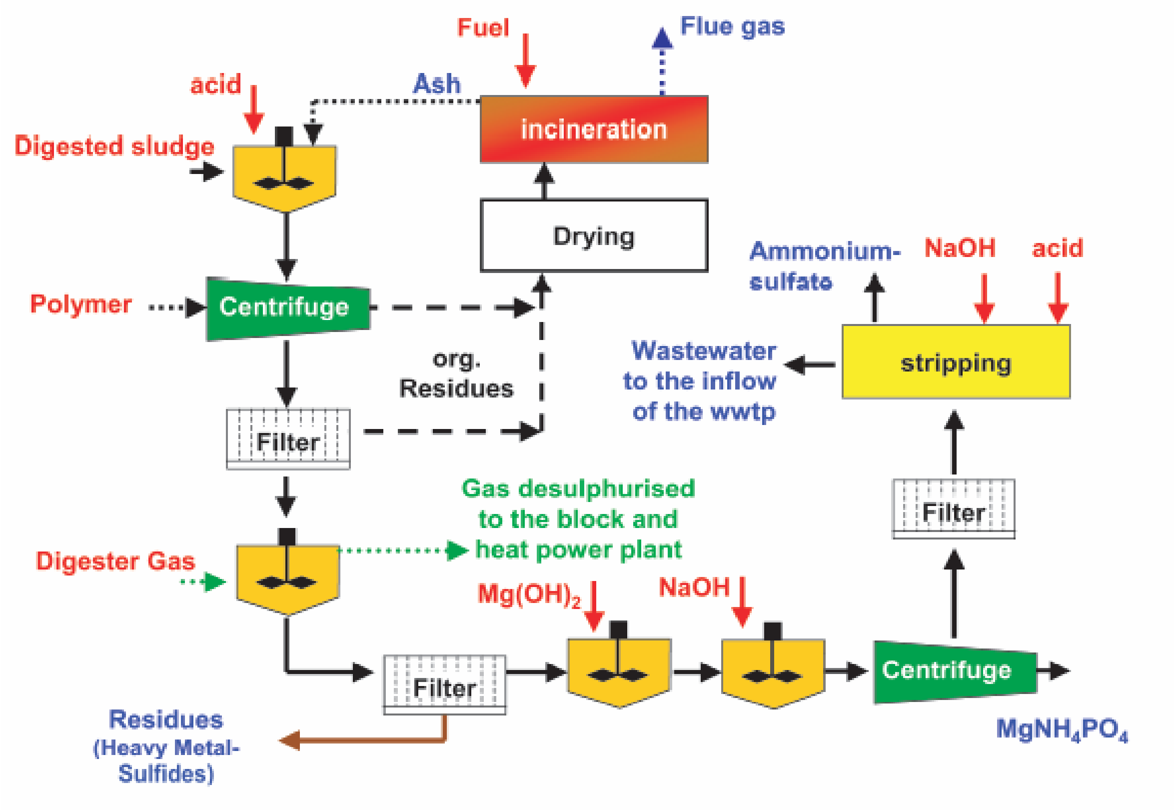

Figure C-3. **Process flowsheet of the Seaborne Technology at the Gifhorn wastewater treatment plant. (From Müller et al., 2007).**

Cost Estimate

No capital or operating costs related to the Seaborne process were identified in the literature

C.1.1.4. BioCon Technology

Description

BioCon Technology was developed for recovering phosphorus as phosphoric acid. Ash from sludge incineration is leached with H_2SO_4, and resources including ferric chloride ($FeCl_3$), potassium bisulfate ($KHSO_4$) and phosphoric acid (H_3PO_4) are recovered by ion exchange (Hultman, et al., 2003; Levlin et al., 2004). The recovery process is indicated in Figure C-4.

Application

BioCon Technology has not yet been investigated at full-scale. However, it has been studied at a pilot-scale at a MWTP near Aalborg in Denmark (Hultman et al., 2001). A full-scale plant was anticipated for the city of Falun, Sweden, but the project was abandoned (Hultman et al., 2003).

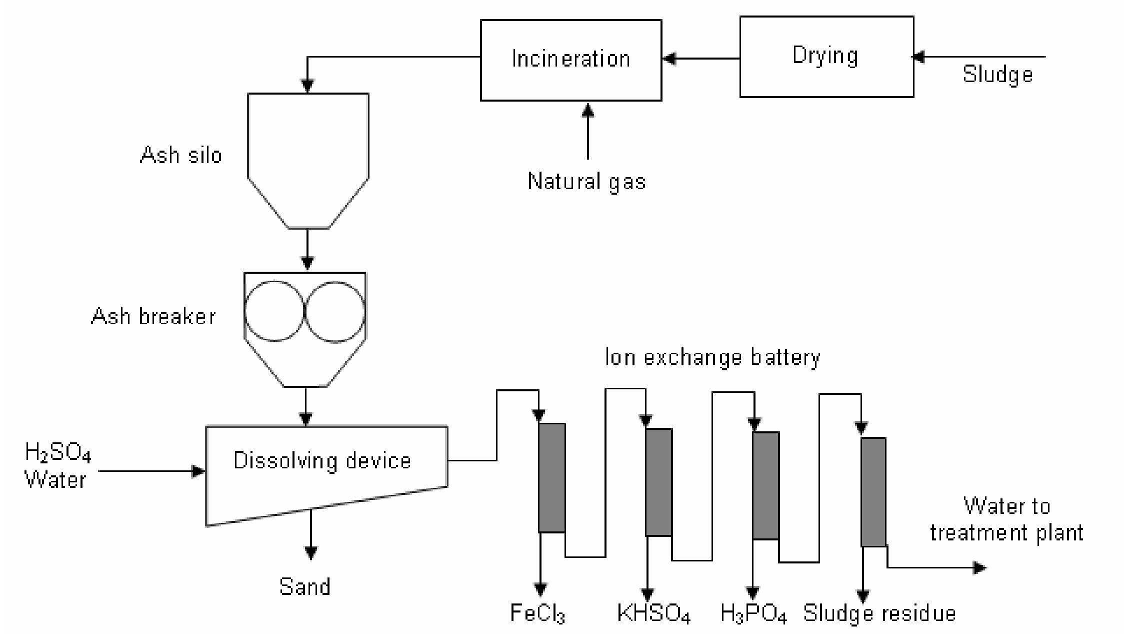

Figure C-4. Flow diagram of the BioCon Technology (From Hultman et al., 2001).

C.1.1.5 SEPHOS Technology

Description

SEPHOS is an acronym for the SEquential precipitation of PHOSphorus. This technology concept has been initiated in Germany. It is under development at the Institute WAR in Germany (Berg and Shaum, 2005). The inventors expect that the SEPHOS Technology will produce aluminium phosphate and calcium phosphate using ash from sludge incineration. The first product could be used in the electrochemical phosphate industry while the second could be reused as fertilizer.

In the first stage (Figure C-5), the ash is mixed with sulfuric acid to decrease the pH to 1.5. After separating the solids from the liquid phase, the pH of the liquid is re-adjusted to around 3.5. At this pH, phosphorus and aluminum present in the liquid precipitate as AlPO4. Copper and zinc only precipitate at pH higher than 3.5 (Schaum et al., 2007).

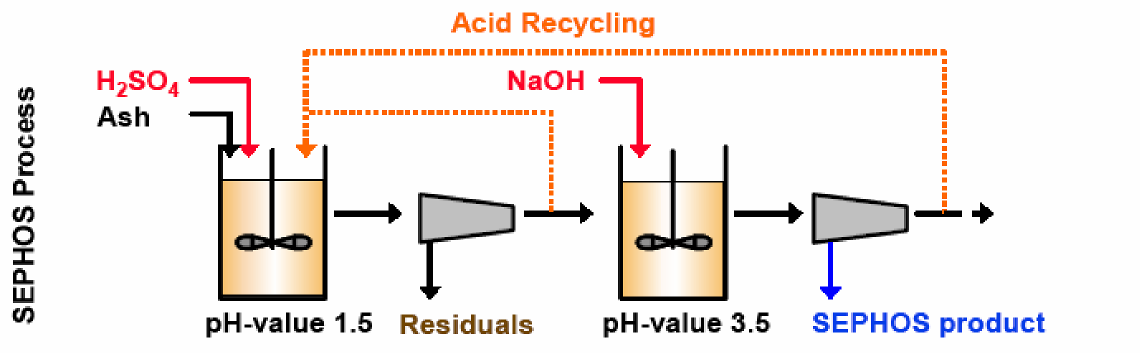

Figure C-5. Flow diagram of the SEPHOS Technology for AlPO4 recovery (From Schaum et al., 2005).

In a second stage (Figure C-6), aluminum phosphate can be dissolved by elution in a basic solution with high pH (12-14). The dissolved phosphorus can be precipitated as calcium phosphate.

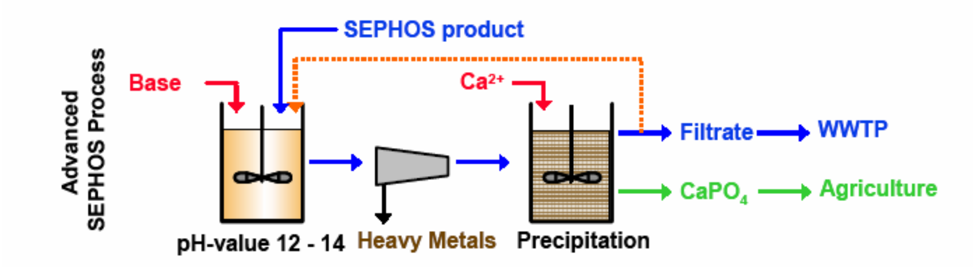

Figure C-6. Flow diagram of the SEPHOS Technology for $CaPO_4$ recovery (From Schaum et al., 2005).

Application

No full-scale or pilot scale demonstration of the SEPHOS technology has been conducted yet. The process looks attractive though since no energy is required for heating.

Cost Estimate

No capital or operating costs related to the Seaborne process were identified in the literature. The current estimated costs for chemical needed by the SEPHOS process are approximately three times higher than the world price of raw phosphate (Schaum et al., 2007).

C.1.1.6 ARP Technology for Nitrogen Recovery

Description

ARP is an acronym for the Ammonia Recovery Process. This technology was developed by the Battelle Memorial Institute. It is currently commercialized by ThermoEnergy.

ARP Technology is a reversible chemo-sorption process. It uses a proprietary resin system to extract a commercial-grade fertilizer from a nitrogen-rich stream. The technology was initially developed to treat the nitrogen-rich effluent after sludge processing with the Sludge-To-Oil Reactor System (STORS). The STORS is an energy recovery process that will be described latter in this Appendix.

In the ARP process, an ion exchange unit concentrates the ammonia in the influent from approximately 1000 ppm to 15,000 ppm (CERF, 2000). The concentrated ammonia stream is then vaporized. Finally the ammonia gas is crystallized as ammonium sulfate. The main steps involved in the ARP processes are shown in Figure C-7.

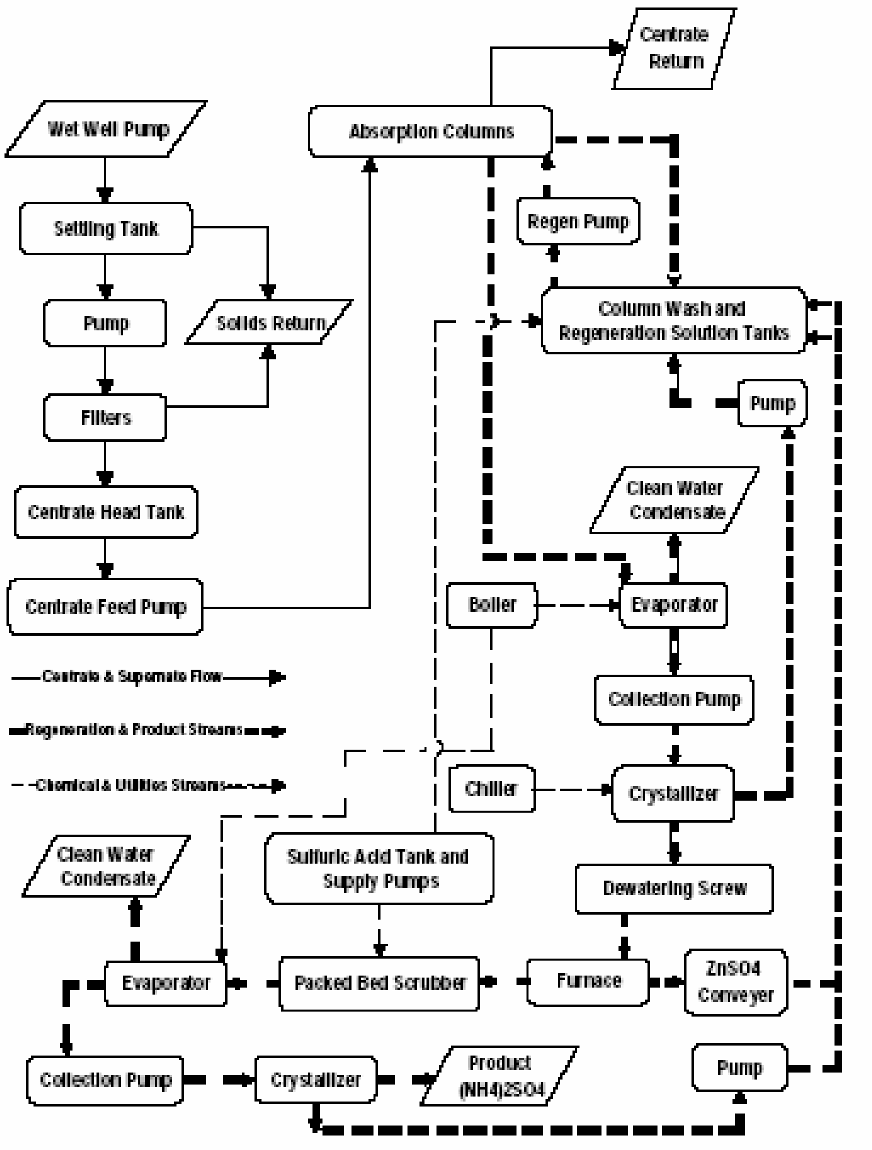

Figure C-7. Schematic of the ARP process (From CERF, 2000).

The main chemical reactions involved in the ARP process are described below (CERF, 2000):

♦ *Ammonia equilibrium chemistry on columns*

Ammonia dissociation in water: $NH_4^+ \leftrightarrow NH_3 + H^+$

Ammonia adsorption onto column resin: $R\text{-}Zn^{2+} + 2HN_3 \leftrightarrow RZn^{2+}{:}(NH_3)_2$

♦ *Column regeneration with sulfuric acid and zinc sulfate*

Ammonia stripping: $RZn^{2+}{:}(NH_3)_2 + H_2SO_4 \leftrightarrow R\text{-}Zn^{2+} + 2NH_4^+ + SO_4^{2-}$

Zinc adsorption equilibrium: $R\text{-}2H^+ + ZnSO_4 \leftrightarrow R\text{-}Zn^{2+} + 2H^+ + SO_4^{2-}$

♦ *Zinc chemistry*

Precipitation of ammonia: $2NH_4^+ + Zn^{2+} + 2SO_4^{2-} + 6H_2O \leftrightarrow (NH_4)_2SO_4ZnSO_4{}^{\bullet}6H_2O_{(s)}$

- *Separating and recovering the ammonia and zinc*

$(NH_4)_2SO_4ZnSO_4{\bullet}6H_2O_{(s)} + \text{heat} \leftrightarrow 2NH_{3(g)} + ZnSO_{4(s)} + SO_{3(g)} + 7H_2O_{(g)}$

$SO_{3(g)} + H_2O_{(l)} \rightarrow H_2SO_{4(l)}$

$2NH_{3(g)} + H_2SO_{4(l)} \rightarrow (NH_4)_2SO_{4(l)}$

Application

The first ARP pilot plant was constructed at Oakwood Beach Water Pollution Control Plant (WPCP) Staten Island, New York. The pilot plant was tested from September through December of 1998. The evaluation was conducted using centrate produced during dewatering operations of the plant's anaerobically digested sludge. The main objectives were to test the process efficiency and estimate the cost for treating a centrate stream.

Cost Estimate

The total treatment cost was evaluated between 0.75 and 1.5 US$ per cubic meter of centrate treated (CERF, 2000). This did not take into account the potential resale value of the recovered product.

C.2 Energy Recovery

C.2.1 Sludge to Biogas

C.2.1.1 Anaerobic Digestion - The Bioterminator$^{24/85}$

Bioterminator$^{24/85}$ is a new mesophilic (35°C) anaerobic digestion technology developed by Total Solids Solutions, LLC from research conducted at the University of Louisiana, U.S. Figure C-8 shows the simplified flow diagram of the process. The main difference between this technology and the conventional anaerobic digestion process is the mixing pattern.

Bioterminator$^{24/85}$ works with laminar plug flow hydraulics as opposed to complete mixing (TA, 2004). This technology is apparently capable of destroying 85% of TS in 24 hours. The detention time of the reactor is 24 hours or less. This system works on secondary or mixed primary/secondary sludge with a solids concentration of 0.5-5.0% TS.

During the digestion process with BioTerminator $^{24/85}$, a supplemental buffering agent may be fed to maintain the pH. This depends on the characteristics of the raw sludge. Additionally, a low dose of sucrose is fed to the digester as a catalyst. The amount of sucrose solution (with 10% sucrose) needed per 80 m^3 of sludge treated is about 1 L. A maximum of 25% of the digester gas is required for heating to maintain the temperature of the digester to 35°C.

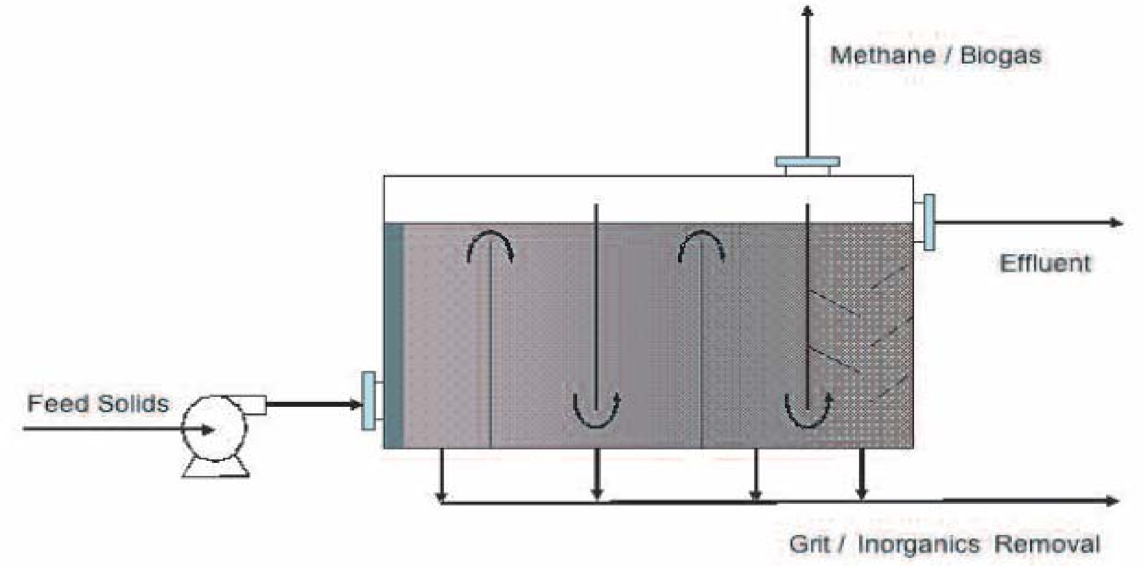

Figure C-8. **Simplified schematic of the Bioterminator24/85 (From Burnett & Togna, 2007).**

According to Burnett and Togna (2007), an unheated pilot scale treating primary sludge achieved an average VS removal efficiency of 93% at a two-day hydraulic retention time. The pilot, 3.785 m^3, was installed at the central MWTP in Baton Rouge, Louisiana, US in October 2000. It operated for five months. The first full-size installation is scheduled to be constructed in Daphne, Alabama in 2007 (U.S. EPA, 2006).

C.2.1.2 Ozonation

Ozonation is an oxidative pre-treatment process that is used to enhance sludge hydrolysis and improve solids biodegradability. This process is of major interest to Kurita Water Industries. Goel et al. (2004) investigated a pilot-scale experiment in Japan. The pilot was built in a full-scale sewage treatment plant. During this study sludge was pre-treated by ozonation and then treated in an anaerobic digester.

The anaerobic digester had a volume of 1.1 m^3. A mixture of municipal sludge, consisting of primary and secondary sludge in the ratio of 1:3.5 (weight by weight) was used during this study. An average ozone dose of 0.026 kgO_3 per kg total volatile solids (TVS) was applied. The pilot treated 20 liters of sludge per day with a volumetric loading rate of 0.51 kg $TVS/m^3.d$.

During the study, part of the digested sludge (70 L/d) was withdrawn from the digester, ozonated and returned back to the reactor. The sludge retention time of this system was very long (275 days) because of the recycle stream. A long retention time is one the key points of the process (Goel, 2007 - personal communication). The longer the ozonated sludge remains in the system better is the removal efficiency.

This new process scheme resulted in organic solid degradation of 81% while the TS degradation efficiency was 61%. The process produced 36% more energy than a digester treating the same quantity and same sludge. However, the production of ozone requires an energy input. The required energy during this experiment was higher than the surplus generated.

The energy input (ozonation plus pumping) and the energy produced were estimated at 1923 kWh and 1736 kWh per dry MT treated, respectively (Goel et al., 2004). This pilot-scale experiment showed that ozone treatment improves the energy production of the anaerobic digester but it has a negative energy balance. The amount of ozone used could be a key point to improve the energy balance of the process.

C.2.2 Sludge-To-Oil

C.2.2.1 Sludge-To-Oil Reactor System

Description

Sludge-To-Oil Reactor System (STORS) is a hydrothermal process. This technology was developed by Battelle Memorial Institute. Preliminary researches were conducted around 1986 and the results were reported by (Molton et al. 1986).

Wet sludge with about 20% dry solids is mixed with 5% anhydrous sodium carbonate (Na_2CO_3) prior to treatment. The reactor operating temperature, pressure and detention time are roughly 275 to 315^{o}C, 11,400 to 14,800 kPa and 1 to 3 hours respectively.

During the treatment, the sludge is converted into a fuel consisting of an oil with 90% of the heating value of diesel, and a solid "char" similar to coal. This material can be used to produce electricity and/or heat using an engine.

Application

The technology is currently commercialized by ThermoEnergy (ThermoEnergy, 2007). There is currently no full-scale installation in operation. The STORS technology has however been applied at pilot scale in California at the Colton MWTP (20,000 m^3/d). During this project, the reactor was combined with the ARP (previously described in this Appendix) to recover both biofuel and ammonia. The pilot operated at a temperature of 315^{o}C. The Colton project began in September 1998 and was concluded in November 2000. The project cost, in 1998 US$, was estimated at 3 million and was funded by the U.S. EPA. The next step for this project is to upgrade the pilot plant to a full-scale commercial plant.

Cost Estimate

Information such as energy input, energy generated, O&M costs for the Colton project is not available in the literature. A general cost estimate of the STORS technology was provided by Molton et al (1986) for three plants sizes including small, medium and large size population. The estimations were based on a prototype sludge-to-oil continuous reactor system. Table C-3 summarizes the main figures, converted in 2007 US$, of their estimations. The original costs are shown between brackets. The costs per dry MT treated decreases significantly as the size of the population serviced increases.

An estimation of energy input and output was performed based on the results reported by Molton et al. (1986) for the prototype system. The results of this estimation are summarized in Table C-3. With regard to the oil generated, the initial numbers in kcal were converted in kWh of electricity. A conversion efficiency of oil to electricity equal to 38% was assumed according to Bridle (2004).

Table C-3. Cost estimate (in 2007 US$) and energy for three facility size (Adapted after Molton et al. 1986).

Parameters	**Town** **10,000 P[a]**	**City** **100,000 P**	**Metropolis** **1 Million P**
Sludge Treated (dry MT per year)	292	2920	29200
Capital cost (US$)	1,278,481 (687000)[c]	2,892,866 (1554500)	11,382,577 (6116500)
Capital cost (US$ per dry MT per year)	*4,378*	*991*	*390*
O&M costs (US$)	104,214 (56000)	241,925 (130000)	1,116,578 (600000)
O&M costs (US$ per dry MT)	*357*	*83*	*38*
Energy used in the process (kWh)[b]	411,667	4,070,558	40,705,580
Energy used in the process (kWh per dry MT)	*1410*	*1394*	*1394*
Electricity from oil produced (kWh)	554,167	4,322,503	43,225,030
Electricity from oil produced (kWh per dry MT)	*1898*	*1480*	*1480*

[a] P indicates persons; [b] Energy to heat the reactor to a temperature of 150°C; [c] Costs in parentheses are the original costs from the authors (in 1986 US$)

C.2.3 Sludge to Liquid

C.2.3.1 Aqua Citrox® Technology

Description

The Aqua Citrox® process operates at a temperature of 400-600°C and a pressure of 25,000 kPa (250 bars). Figure C-9 shows the process flowsheet of the process. After pre-treating the sludge with a macerator to remove large particles, the feed passes through a high pressure pump to increase the pressure to 25,000 kPa (250 bars). It is then heated before it enters in the SCWO reactor. The heat used for this operation is provided by the effluent from the reactor. The detention time of the reactor is about 1 minute. The Aqua Citrox® process is capable of treating a stream of at least 15% of dry solids. The process uses a high temperature in order to destroy completely the nitrogen contained in the sludge.

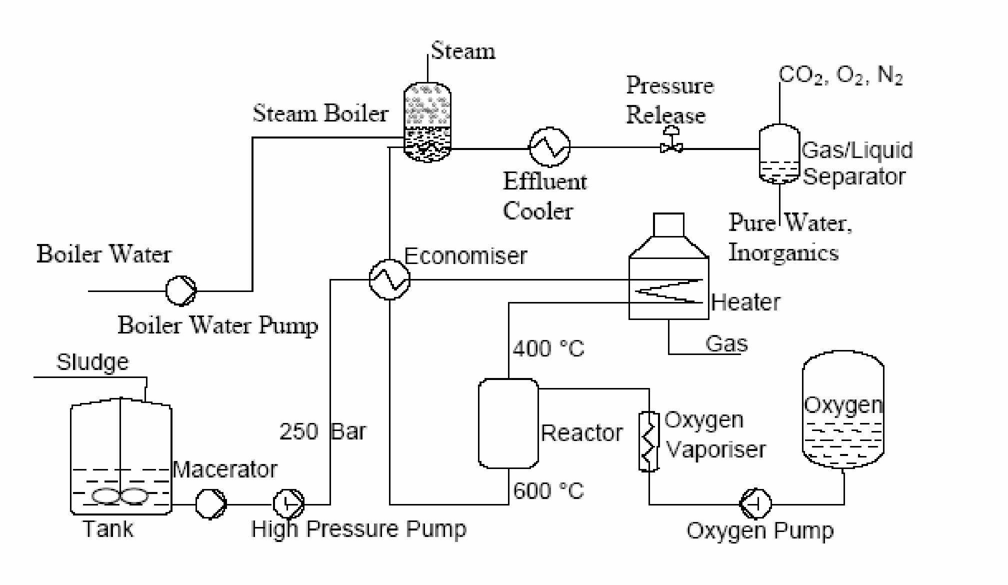

Figure C-9. Process diagram of the Aqua Citrox® Technology (From Gidner & Stenmark, 2001).

Application

The Aqua Citrox® was developed in 1995 by Chematur Engineering in Japan. A pilot scale unit with a capacity of about 1100 kg per hour was built in around 2000 in Japan. This experiment showed that the removal efficiency of COD and nitrogen by the Aqua Citrox® process was higher than 99.99%.

Operation of the Aqua Citrox® requires electricity (240 kWh per dry MT treated), natural gas (23 m^3 per dry MT treated) and oxygen (1100 kg per dry MT treated) as reported by Gidner and Stenmark, (2001). The total energy input (electricity, natural gas and oxygen production) can therefore be estimated at approximately 1576 kWh per dry MT. Again, this assumes that the production of 2 tonnes of oxygen requires 1 MWh of electricity. The amount of energy that could be recovered is not available.

Cost Estimate

Gidner and Stenmark, (2001) estimated that the capital cost of the installation per dry MT of sludge treated is about 79 GBP while the O&M costs are about 70 GBP per dry MT. These costs are approximately equivalent to US$158 and US$140 per year, respectively. The capital cost was estimated assuming an interest rate of 8% and a depreciation time of 12 years. This cost includes only the SCWO unit, the oxygen system, the building and auxiliary equipments such tanks. It does not include the cost for the sludge concentration unit (dewatering).

C.3 ENERGY AND RESOURCE RECOVERY

C.3.1 KTH Two-Stage Acid-Base Leaching Concept

Description

Discussion above has reported on either energy recovery or resource recovery, but not an integration of both concepts. A two step process concept using both acid and base extractions of phosphorus was suggested by Kungl Tekniska Högskolan (KTH), the Swedish Royal Institute of Technology (Levlin and Hultman, 2004). The KTH concept for energy recovery and resource recovery is shown in Figure C-10. The concept is based on recovery of products after two different stages (Hultman, 1999). Sludge from a BNR system is treated in an anaerobic digester to, produce biogas (energy) and release phosphorus and ammonium (resources). The digested sludge solids are separated from the liquid phase (supernatant). The latter is used to recover struvite or to produce phosphoric acid by extraction. The digested sludge can be conditioned with heat, pressure, acids, etc. and dewatered. The filtrate from dewatering is further treated to selectively remove toxic metals. The remaining sludge may be used for additional energy recovery (e.g. incineration).

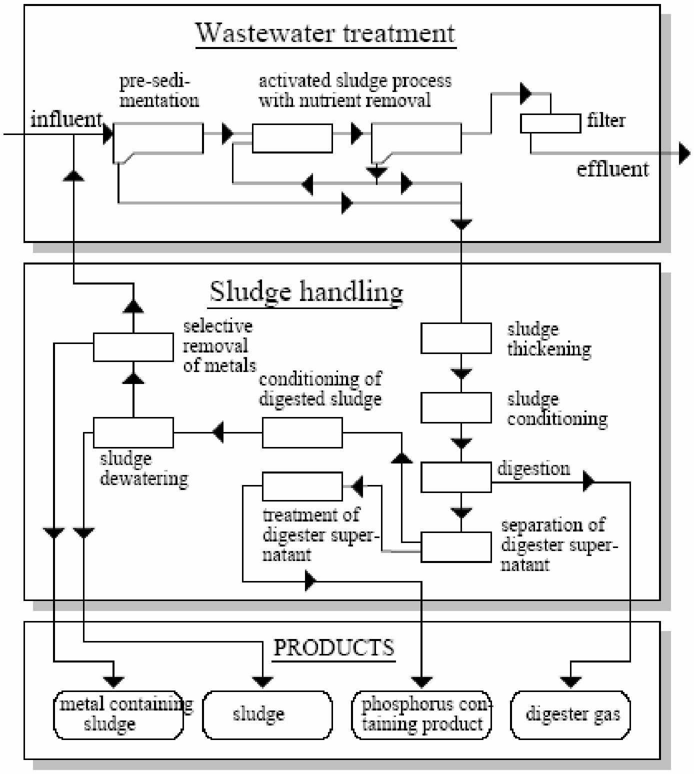

Figure C-10. Sludge handling for Product Recovery (From Hultman, 1999).

Phosphorus recovery from ash or SCWO residue is depicted in Figure C-11. The first step uses a low dose of HCl to dissolve calcium, magnesium and part of the phosphorus contained in the ash. A low dose is necessary to avoid leaching of Aluminum (Al^{3+}); a high dose of HCl would generate more phosphorus, but it would also dissolve unwanted Al^{3+}. In this case aluminium will have to be separated from the leachate prior recovering the phosphorus.

The leachate from the first step is further treated with H_2SO_4 to produce phosphoric acid (H_3PO_4). Calcium phosphate can then be produced through addition of lime. In the second step of this system, NaOH is used to leach the remaining phosphorus in the solids from the first stage.

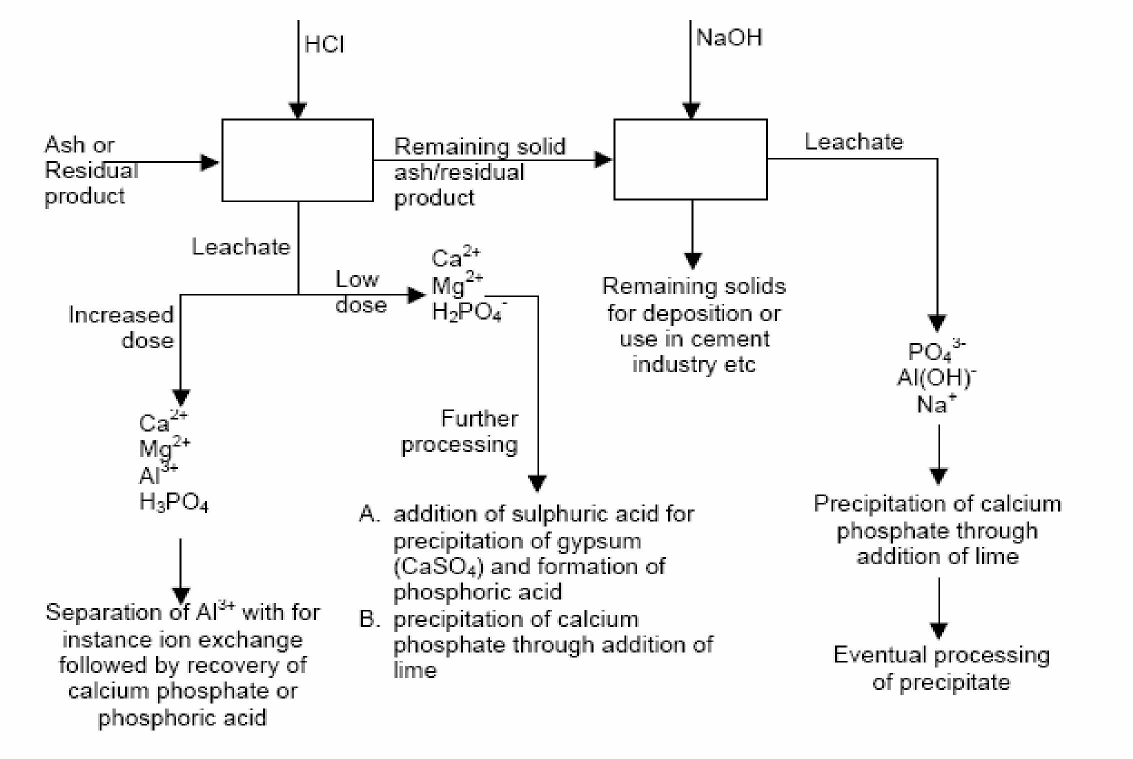

Figure C-11. **System for phosphorus recovery from ash and SCWO-residual product (From Levlin & Hultman, 2004).**

Application

The process has been tested at bench scale (Levlin, 2006; Levlin, 2007). Based on the results obtained, the cost of chemicals for two-step leaching is estimated about €1.2 per kg of phosphorus removed (Levlin, 2006). No full-scale application has been reported to date. The anticipated advantage of the process is a low heavy metal contamination of the final product. A similar concept named SEPHOS, under development in Germany, was described earlier in the report.

C.3.2 Aqua-Reci^{TM} Technology

Description – Resource Recovery

The Aqua Reci^{TM} Technology was developed in Sweden jointly by Chematur Engineering AB and Feralco AB (Chematur, 2007). Figure C-12 shows the flow diagram of the Aqua Reci Technology. The process recovers both a resource material (phosphorus) and energy. This technology combines the SCWO mechanism with an extraction method. The SCWO process, described earlier in this Appendix, is used to produce ash. Then phosphorus is extracted from the ash.

Phosphorus is extracted by leaching the ash with HCl, H_2SO_4 or NaOH. Leaching the ash with 1% NaOH solution at a temperature between 80-90^{o}C during 75 to 90 minutes, can extract 65% to 90% of the phosphorus from the ash (Stendahl and Jäfverström, 2004). Extraction efficiencies are higher with acid. About 100% of the phosphorus could be extracted with HCl or H_2SO_4 at a temperature of 90^{o}C during about 2 hours.

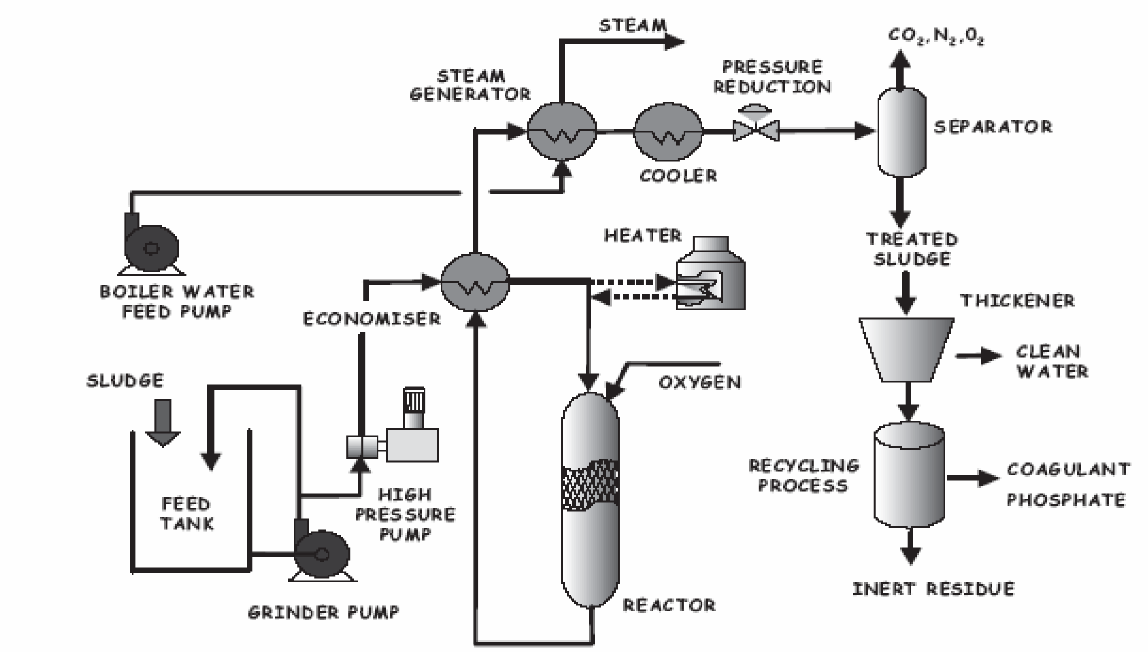

Figure C-12. Aqua Reci™ Technology (From Stendahl & Jäfverström, 2004).

Levlin et al. (2004) and Stark (2005) in Sweden conducted leaching and recovery experiments on residue from SCWO and ash from incineration. The experiments were performed at laboratory scale using HCl and NaOH.

The results of their experiments are summarised in Table C-4. The results show that at low acid concentration, it is easier to release phosphate from the SCWO residue than from the incineration ash. Apparently, one of the advantages of SCWO is the quality of residue that is generated. According to the authors, the residue from SCWO is very fine, thus allowing a better contact with the extraction solution. It is important to note that the two sludge substrates were from different MWTPs. The SCWO residues were obtained from Broma, Borlänge and Karlskoga while the incineration ash was from Mora.

Table C-4 shows as well that, phosphate was leached more easily with HCl than NaOH at room temperature. This was also observed by Stendahl and Jäfverström (2004). Although acid leaching is more efficient, alkaline leaching gives lower metal contamination in the leachate (Stark, 2005; Levlin et al., 2004). Without an appropriate method for separating phosphorus from metal contaminants, phosphorus recovered from extraction with HCl might become a problem for agricultural application with stringent application conditions.

The experiments revealed as well that increasing the extraction temperature from 20°C to 90°C during chemical treatment seems to have no effect on phosphate leaching. This indicates that technologies based on chemical leaching for phosphate extraction should not be energy intensive.

Table C-4. Experiment conditions and performance data – Acid leaching versus Alkaline leaching (Stark, 2005; Levlin et al., 2004).

Parameter	Acid leaching with HCl		Alkaline leaching with NaOH	
Residue	SCWO residue	Incineration Ash	SCWO residue	Incineration Ash
Amount of Residue	5 mL	0.5 g	5 mL	0.5 g
Chemical addition (mL)	25	25	25	25
Chemical concentration (M)	0.1	1	1	1
Temperature (^{o}C)	20	20	20	20
Chemical contact time (h)	2	2	2	2
4000 rpm centrifuge (min)[a]	20	20	20	20
Efficiency (%)	80-100	75-90	50-70	40-70

[a] Centrifuge time, after chemical contact, for leachate-solid separation

Demonstration of the technology at full-scale would improve the potential for phosphorus recovery, due to the low chemical requirement and high extraction efficiency with SCWO residues. In addition, the SCWO process operates at a lower temperature than incineration, making the SCWO less demanding in term of energy input. While electricity can not be recovered with SCWO, the ability to recover phosphorus may improve the feasibility of the technology.

<u>Description – Energy Recovery</u>

Based on the pilot-scale experiment of Karlskoga, Sweden, a material and energy balance of the process was estimated by Stendahl and Jäfverström (2003). Their estimation is depicted in Figure C-13.

A system treating 128,000 MT of waste (15% dry solids) per year will need 5,760 MWh of electricity and 16,800 tonnes of oxygen. The plant will then generate 60,800 MWh of energy, thus an excess energy could be recovered. On this basis, the plant needs 300 kWh per dry MT and produces 3,167 kWh per dry MT.

Assuming again here that the production of 2 tonnes of oxygen requires 1 MWh of electricity, 438 kWh per dry MT are required for the production of 16,800 tonnes of oxygen. The process total energy input is therefore about 738 kWh per dry MT. Nevertheless, the process still a large excess energy (heat) producer, about 2,429 kWh per dry MT.

The estimation indicates that only 3% of the initial amount of feed should leave the process as waste. If the results were confirmed with full-scale application, this process would be a very attractive technology for sludge treatment provided that the cost is affordable by municipalities.

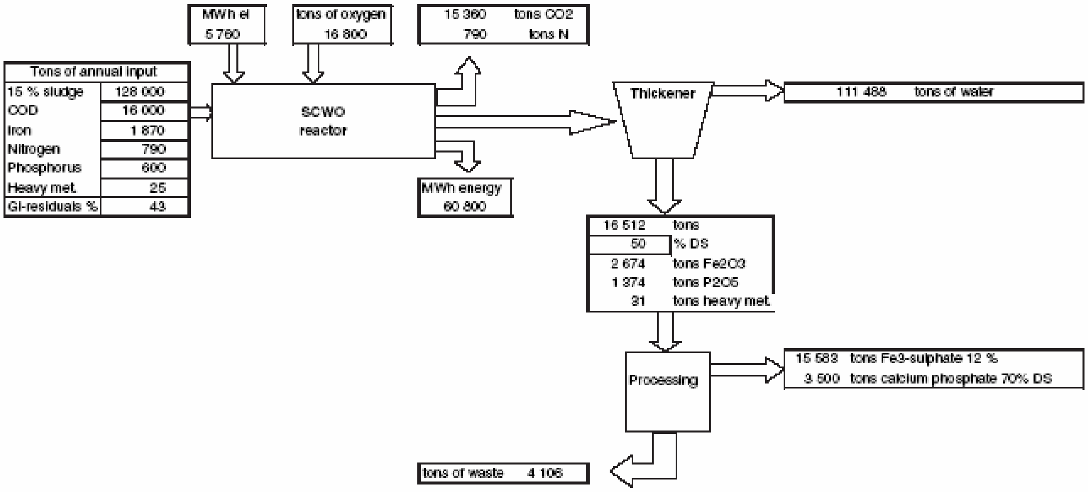

Figure C-13. Material and energy balance for the Aqua Reci^{TM} Process (From Stendahl & Jäfverström, 2003).

Application

A pilot scale Aqua Reci^{TM} Technology was tested to recover phosphorus in Karlskoga, Sweden. The test was conducted with digested sludge (15% dry solids) from sewage treatment plants in Stockholm and from Karlskoga sewage treatment plant.

Cost Estimate

The cost for a SCWO system treating 10 dry tones per day of sewage sludge, reported by Svanström et al. (2004), is presented in Table C-5. The cost does not include the unit for phosphorus recovery. This estimate assumes that liquid CO_2 is recovered and CO_2 credit is applicable. Based on the data in Table C-5 it appears that the capital and O&M costs when considering energy alone are in the order of US$740 per dry MT per year and US$ 230 per year, respectively.

Stendahl and Jäfverström (2003) estimated the total investment for the phosphorus recovery system to cover the total demand of the Stockholm sewage works at €660,000 (about US$ 990,000. Based on this information, the total cost of the Aqua Reci^{TM} process for the full-scale Stockholm plant is approximately US$ 946 per dry MT treated per year.

Table C-5. Cost breakdown for a 10 dry MT treated per year with SCWO system (Svanström et al., 2004).

Parameter	**Value (US$)**
Installed Capital Cost	2700000
Annualized Costs	
Cost of capital (10%, 10 yr)	396000
Maintenance (10% of major equipment cost)	79000
Operating direct labour	352000
Oxygen	350000
Electricity	19000
Cooling water	16000
Residual solids disposal	22000
Annual Operating Cost	1234000
Annualized Credits	
Hot water production	(264000)
Carbon dioxide by-product	(119000)
Annual Credit	(383000)
Annual Net Cost	851000
Unit Cost (US$/dry ton)	243

Appendix D:

Conversion Factors

Table D-1. Unit Conversions.

Convert From	To	Multiply by	Cautions and Assumptions
Energy			
British thermal unit (BTU)	Mega joule (MJ)	0.001054	Assuming thermal BTU
Kilo joule (kJ)	MBTU	0.000948452	Assuming thermal BTU
Kilowatt hour (kWh)	Mega joule (MJ)	3.6	
Mega joule (MJ)	Kilowatt hour (kWh)	0.277778	
Mega joule (MJ)	British thermal unit (BTU)	948.451653	
MBTU	Kilo joule (kJ)	1054.35	
MBTU	British thermal unit (BTU)	1000	
MMBTU	British thermal unit (BTU)	1000000	
Length			
Foot (ft)	Meter (m)	0.3048	
Inch	Millimeter (mm)	25.4	
Meter (m)	Foot (ft)	3.28084	
Millimeter (mm)	Inch	0.0393708	
Mass			
Kilogram (kg)	Pound (lb)	2.204622	
Kilogram (kg)	Short (ton)	0.001102311	Assume US (or short) ton
Kilogram (kg)	Metric tonne (MT)	0.001	
Mega gram (Mg)	Metric tonne (MT)	1	
Metric tonne (MT)	Kilogram (kg)	1,000	
Pound (lb)	Kilogram (kg)	0.4535924	
Short ton (ton)	Kilogram (kg)	907.1847	
Mass per volume			
Kilogram per cubic meter (kg/m^3)	Pound per cubic foot (lb/ft^3)	0.062428	
Kilogram per cubic meter (kg/m^3)	mg/L	1,000	
Kilogram per cubic meter (kg/m^3)	Percentage (%) solids	0.1	
Milligram per liter (mg/L)	Kilogram per cubic meter (kg/m^3)	0.001	
Milligram per liter (mg/L)	Percentage (%) solids	0.0001	
Percentage (%) solids	Kilogram per cubic meter (kg/m^3)	10	
Percentage (%) solids	Milligram per liter (mg/L)	10,000	
Pound per cubic foot (lb/ft^3)	Kilogram per cubic meter (kg/m^3)	16.01846	

Table D-1. Unit Conversions (continued)

Convert From	To	Multiply by	Cautions and Assumptions
Power			
Horse power (hp)	Kilo watt (kW)	0.746	
Joule per second (J/s)	Kilo watt (kW)	0.001	
Kilo watt (kW)	Horst power (hp)	1.34048	Assume motor/electric/ international hp
Kilo watt (kW)	Joule per second (J/s)	1000	
Kilo watt (kW)	Mega joule per day (MJ/day)	86.4	
Mega joule per day (MJ/day)	Kilo watt (kW)	0.011574	
bhp(brake hp) = net available power(already includes engine efficiency)			
ie. bhp (brake hp) = hp x efficiency			
Power - Boilers Only			
Boiler horst power (bhp)	British thermal unit per hour (BTU/hr)	33472.119	
Boiler horst power (bhp)	hp (electric)	13.15	
Boiler horst power (bhp)	Kilo watt (kW)	9.8097	not same as brake hp
Boiler horst power (bhp)	Mega joule per day (MJ/day)	847.6312	
British thermal unit per hour (BTU/hr)	Kilo watt (kW)	0.00029307	
British thermal unit per hour (BTU/hr)	Mega joule per day (MJ/day)	0.02532	
Kilo watt (kW)	British thermal unit per hour (BTU/hr)	3412.14245	
Temperature			
Degree Celsius (°C)	Degree Fahrenheit (°F)	1.8(°C) + 32	
Degree Fahrenheit (°F)	Degree Celsius (°C)	(°F -32)/1.8	
Volume			
Cubic foot (ft^3)	Cubic meter (m^3)	0.02831685	
Cubic foot (ft^3) Natural Gas	British thermal unit (BTU)	1,000	
Cubic meter (m^3)	Cubic foot (ft^3)	35.31466	
Cubic meter (m^3)	Gallon (gal)	264.172	Assume US liquid gallon
Cubic meter (m^3)	Liter (L)	1,000	
Gallon (gal)	Cubic meter (m^3)	0.003785	
Liter (L)	Cubic meter (m^3)	0.001	
Mcf = Mscf = kscf	Cubic foot (ft^3)	1,000	scf = standard ft^3
MMcf = MMscf	Cubic foot (ft^3)	1,000,000	
10 therms	British thermal unit (BTU)	1,000,000	
1 therm	Cubic foot (ft^3) Natural Gas	100	approximate

Table D-1. Unit Conversions (continued)

Convert From	To	Multiply by	Cautions and Assumptions
Volumetric Flows			
Cubic foot per day (ft^3/d)	Cubic meter per day (m^3/d)	0.02831685	
Cubic meter per day (m^3/d)	Cubic foot per day (ft^3/d)	35.31466	
Cubic meter per day (m^3/d)	Gallon per day (gpd)	264.172	Assume US liquid gallon
Cubic meter per day (m^3/d)	Mega gallon per day (Mgd)	0.0002642	
Gallon per day (gpd)	Cubic meter per day (m^3/d)	0.003785	
Mega gallon per day (MGD) or 1,000,000 gpd	Cubic meter per day (m^3/d)	3785	

Digital Dutch, 2005
Estes, 1981
NW Natural, 2005
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