

**Pilot-scale anaerobic digestion of municipal biowaste with
thermal hydrolysis pre-treatment**

YINGJUN ZHOU

2013

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Chapter 1 Introduction

1.1 Background

Municipal biowaste (MBW), which means the biodegradable wastes produced from municipal facilities, has caused attention worldwide due to its special properties and large production. Compared with other solid waste, MBW usually has higher contents of water and biodegradable organics. These properties cause problems when biowaste are directly disposed with other solid waste, such as production of greenhouse gas and high-strength leachate during landfill, production of dioxin by unsteady burning during incineration, and problem of odour during composting (Hartmann and Ahring, 2006). However at the same time, MBW can be a potential energy resource, which is meaningful due to the present energy crisis. Bio-energy inside MBW can be recovered in the form of methane or hydrogen by anaerobic digestion (De Baere, 2006; DiStefano and Belenky, 2009; El Hanandeh and El-Zein, 2010).

Various methods have been tried to improve the digestion efficiency of MBW. Usually this process is rate-limited by solid hydrolysis. Many physic-chemical methods has been tried to improve hydrolysis rate of anaerobic digestion (Mata-Alvarez et al., 2000). Thermal hydrolysis has been proved one of the useful pre-treatment methods. It's thought that thermal hydrolysis can break cell structure to release organics inside the cells and consequently improve digestion performance (Phothilangka et al., 2008). Anaerobic digestion of sewage sludge with thermal hydrolysis pre-treatment has been applied since 1970s (Fisher and Swanwick, 1971; Haug et al., 1978). Commercial technologies have also been developed based on thermal hydrolysis, known as CambiTM by a Norwegian company (1995) and BioTHELYSTM by a French company (2006). However, co-digestion of WSS with other types of MBW is becoming popular these years as it can improve biogas output and stability of the anaerobic digesters (Hartmann and Ahring, 2006; De Baere, 2006; Bouallagui et al., 2009), while little studies were carried out on the effect of thermal hydrolysis on these mixed MBWs. Qiao et al. (2012) has found that feeding material significantly affected the performance of thermal hydrolysis. As a result, evaluation is needed to clarify the effect of thermal hydrolysis

on different MBWs. Present studies on other MBWs are mainly in lab scale and lack of pilot or large scale operation experience.

On the other hand, digestion efficiency can be improved from the aspect of digester types. The key point is to separate different reaction phases, like separation of different microorganisms (e.g.: two-phase digesters) and substrates (e.g.: anaerobic sequencing batch reactors). It was found thermal hydrolysis can improve settling performance of the biowaste and thus anaerobic sequencing batch reactors can be used for digestion (Wang et al., 2009). Former studies (Dague et al., 1992; Wang et al., 2009) have proved anaerobic sequencing batch reactor may outperform continuously stirred tank reactor in many aspects: 1) better effluent quality; 2) higher digestion efficiency; and 3) higher stability. It needs a smaller hydraulic retention time when same digestion performance is required, which means the construction cost can be reduced with reducing volume. As it can be easily switched from continuously stirred tank reactor without structure variation of the digesters, cost can also be saved for re-construction in a present plant. Operation of ASBR is also simple.

At present, only lab-scale studies have been carried out on the combination of thermal hydrolysis pre-treatment with anaerobic sequencing batch reactors. However, the solid settling behaviours may be totally different from that in a pilot- or large- scale digester. Studies are also lack for modelling of the digestion processes in anaerobic sequencing batch reactors and full evaluation of this new system.

1.2 Objectives

The objectives of this thesis are:

- 1) To clarify the effect of pilot-scale thermal hydrolysis on the properties of municipal biowaste;
- 2) To clarify the effect of thermal hydrolysis on performance of anaerobic digestion in a pilot scale;
- 3) To verify the possibility of using pilot-scale ASBR for the anaerobic digestion of thermally hydrolyzed municipal biowaste and to find the solid settling behaviours in the ASBRs;

- 4) To develop a model from experiment data, which is able to be used for prediction of digester performance, effluent properties, and solid accumulation for both continuously stirred tank reactors and anaerobic sequencing batch reactors;
- 5) To make economic, environmental, and energetic evaluation of the treatment processes used in this study.

1.3 Contents

Structure of this thesis can be described as Figure 1.1.

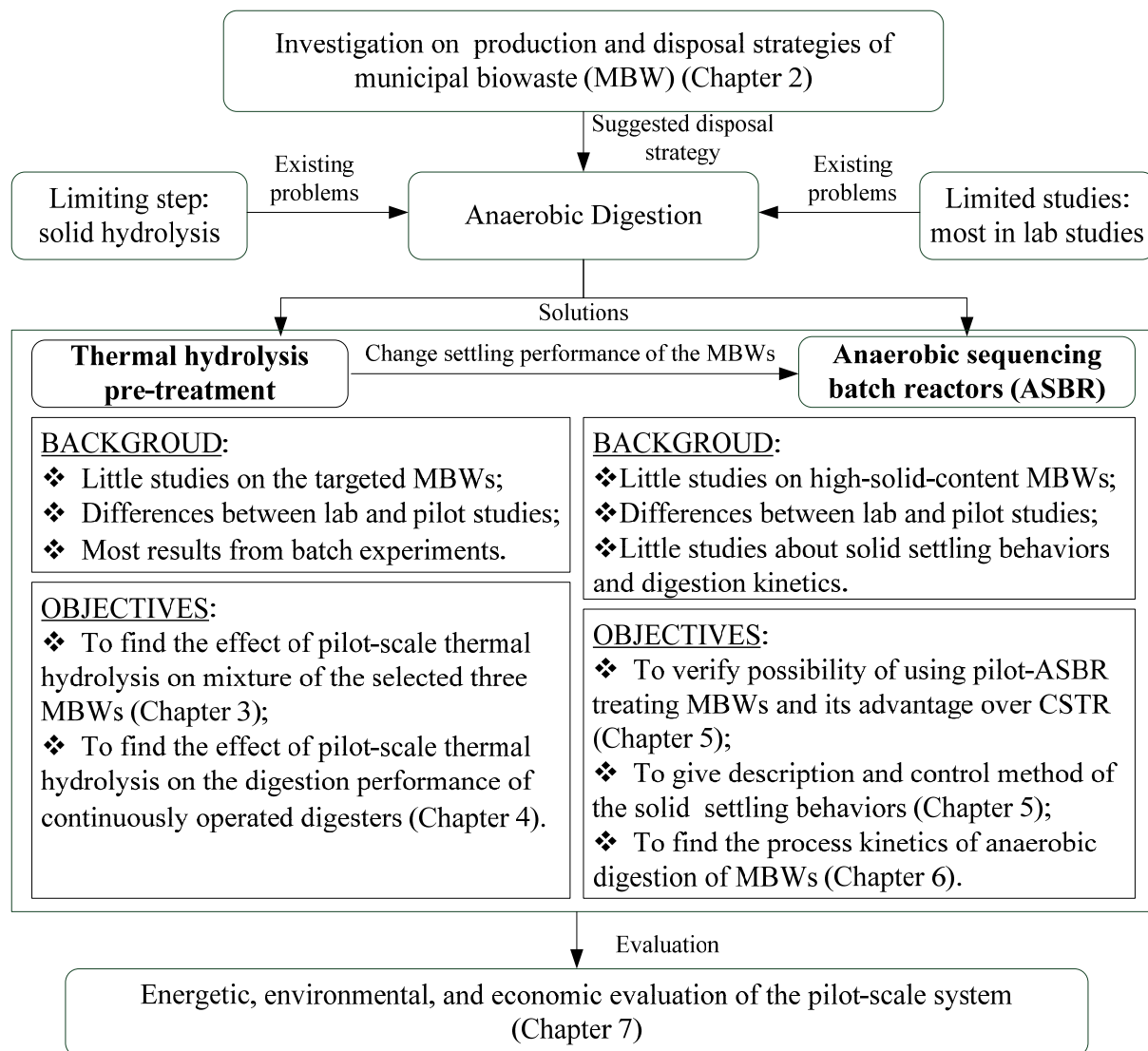


Figure 1.1 Thesis structure

Based on the results of reference survey, it was found application of anaerobic digestion before final disposal could be a promising strategy for the treatment of MBW. However,

anaerobic digestion of MBW is limited by solid hydrolysis. To solve this problem, thermal hydrolysis and ASBR were used.

Thermal hydrolysis was used as pre-treatment to improve solid hydrolysis. As little studies have been carried out on the targeted MBWs and the former relative studies were limited to lab scale, effect of pilot-scale thermal hydrolysis needs to be verified. Moreover, most studies found the effect of thermal hydrolysis from batch experiments, which has some differences from the real digesters which are continuously operated. As a result, it's also needed to check the effect of thermal hydrolysis in the continuously operated digesters.

As thermal hydrolysis improves settling performance of the MBW, ASBR was used as digesters in the anaerobic digestion processes. Solid retention time (SRT) can be divided from hydraulic retention time (HRT) in ASBR, thus solid hydrolysis can be enhanced. However, most studies focused on wastewater treatment, while little cared about treatment of high-solid-content MBWs. The existing studies were limited to lab scale only. Moreover, there're little studies about the settling behaviours and reaction kinetics in the ASBRs. As a result, this thesis studied the possibility to use pilot-ASBR to treat thermally hydrolyzed MBWs. Its performance was compared with continuously stirred tank reactor (CSTR). Solid settling behaviours and reaction kinetics were also described.

Contents of each chapter are as following:

Chapter 1: Introduction

Chapter 2: Reference survey was carried out, mainly focused on production and present disposing methods of municipal biowaste, progress and challenge in anaerobic digestion of municipal biowaste, application of thermal hydrolysis as pre-treatment before anaerobic digestion, and application of anaerobic sequencing batch reactor.

Chapter 3: Effect of thermal hydrolysis on the physical and chemical properties of the municipal biowaste was verified. Results were compared with those from lab studies, which used different heat source from this pilot study.

Chapter 4: Effect of thermal hydrolysis on the anaerobic digestion process was studied. In this chapter, continuously operated pilot-digesters with and without thermal hydrolysis pre-treatment were compared for their performance.

Chapter 5: ASBR was used as digester treating thermally hydrolyzed MBW. Solid settling behaviours were studied in this Chapter and the performance of ASBR was compared with CSTR.

Chapter 6: Modelling and simulation of the anaerobic digestion processes in both ASBR and CSTR. SRT was used as an important parameter in the modelling process to describe reaction kinetics.

Chapter 7: Systemic evaluation was carried out to the energetic, environmental, and economic feasibility of the system.

Chapter 8: Conclusion and suggestions.

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Chapter 2 Reference survey

2.1 Production and disposal of municipal biowaste

Municipal biowaste (MBW) usually refers to the organic fractions of municipal solid waste (OFMSW) (USEPA, 2012a). However in this study, it has been extended to include similar biodegradable waste produced or collected by other municipal facilities. For example, sewage sludge and manure was also taken as MBW. Similar definition was also adopted by the Environment Agency of UK (2012).

2.1.1 Organic fractions of municipal solid waste

Municipal solid waste (MSW) refers to the stream of garbage collected through municipal sanitation facilities. It mainly includes food wastes, yard wastes, containers and product packaging from residential, commercial, institutional, and industrial sources (USEPA, 2012b). With increasing world's population and improved living quality, the production of municipal solid waste (MSW) is increasing in these decades (Figure 2.1).

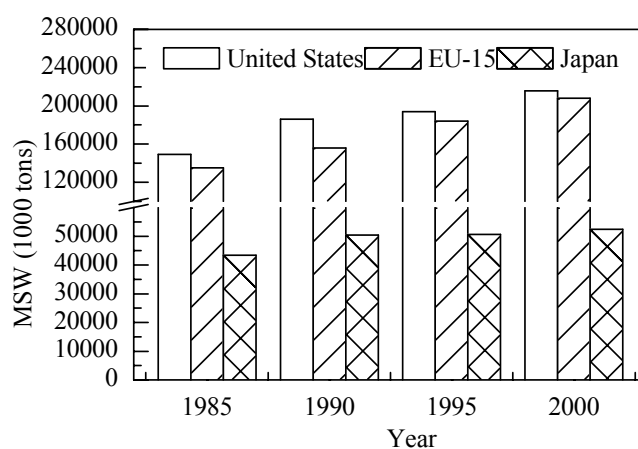


Figure 2.1 MSW production in developed countries (OECD, 2008)

In some areas like Japan, the production of MSW has been decreased recently due to promotion of waste reduce, recycle and reuse (3R) (Figure 2.2). However, MSW production is increasing significantly in those developing countries like China (Figure 2.3), due to fast growth of population and economics in these areas.

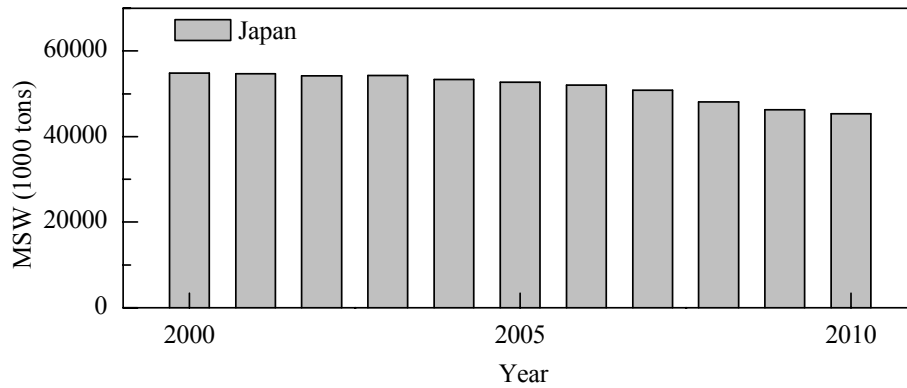


Figure 2.2 MSW production in Japan (Japan Ministry of the Environment, 2012a)

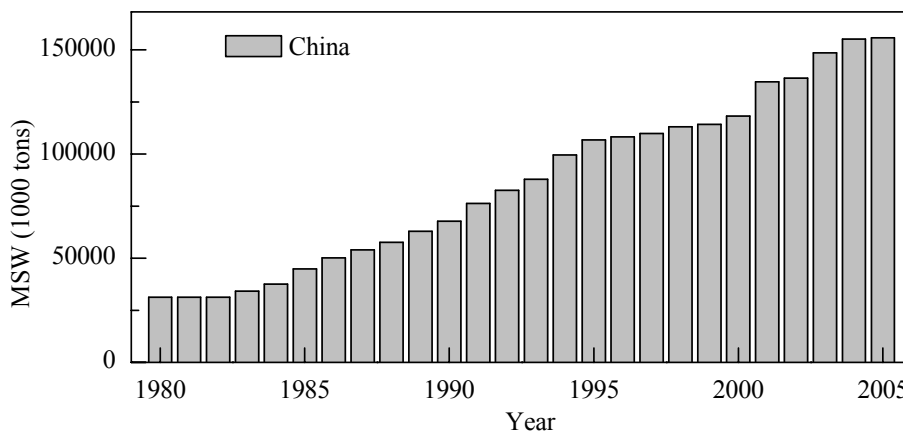


Figure 2.3 MSW collected and transported in China (MOHURD, 1981-2003; National Bureau of Statistic of China, 2012)

Among the MSW, OFMSW takes a large part and has attracted attention for their special properties. It predominantly includes papers, food waste, yard waste, and so on. OFMSW occupies more than 50% of MSW in western countries and Japan (Brummeler and Koster, 1989; OECD, 2008). This proportion is even higher in some developing countries like China. It was reported that 59% of MSW in dry weight was MBW in China in 2002 (Huang et al., 2006). In Beijing, household kitchen waste only occupied 63% of the MSW in 2006 (Li et al., 2009).

2.1.2 Restaurant kitchen waste and market fruit-vegetable waste

In this thesis, two types of OFMSW, together with wasted sewage sludge (WSS), were selected as treatment targets. This two types of OFMSW are restaurant kitchen waste (RKW),

and fruit-vegetable waste (FVW) from food markets. Both of them are produced from relatively concentrated point sources and thus can be collected separately from other MSW.

RKW, together with FVW and household kitchen waste, are recognized as food waste. Figure 2.4 gives MSW composition in Beijing and Kyoto. It can be found food waste takes a large proportion of the MSW production in both cities, which is 55% in Beijing (2005) and 37% in Kyoto (2010). Especially, this proportion kept growing in Beijing these years.

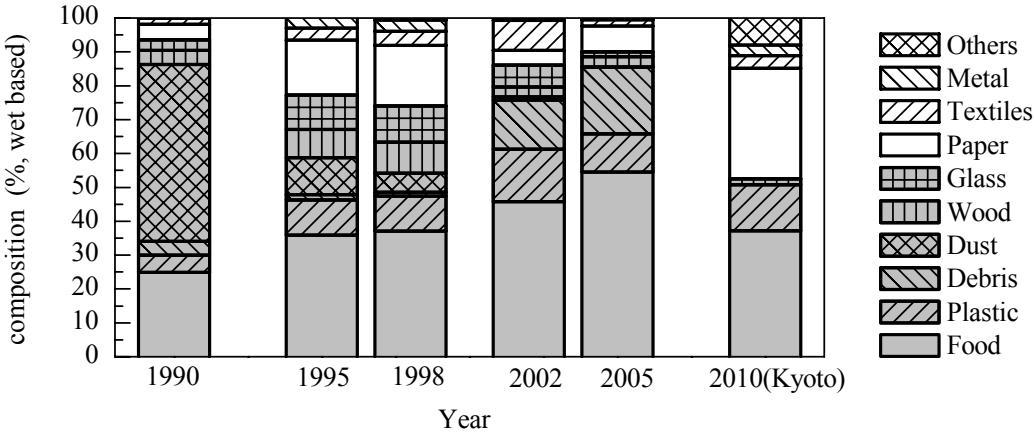


Figure 2.4 Composition of MSW in Beijing and Kyoto (Li, et al., 2009; Kyoto Environmental Bureau, 2012)

It was accounted that RKW takes about 10% of the MSW and exceeds 1000 t/d in China’s megacities like Beijing, Shanghai and Shenzhen. For FVW, a production of 600-1000 t/d was estimated from Beijing’s major food markets. In Shanghai, annual production of FVW reached 500,000 t, which accounted for about 8% of MSW (Hu, 2009).

2.1.3 Wasted sewage sludge

WSS is usually considered as industrial solid waste and is not included in MSW. Figure 2.5 gives yearly production of WSS. It kept growing these years with increase of sewers. It was found 18% of the industrial solid wastes produced in 2003 were WSS, which accounted for 7484 thousand t in wet base (Japan Ministry of the Environment, 2012b).

In China, production of WSS is also increasing quickly these years due to increasing production of sewage and related facilities (as shown in Figure 2.6). By supposing a sludge yield of 0.0005 t dewatered WSS (water content of 80%) per t sewage, it was estimated that about 19 million t of dewatered WSS was produced in 2010.

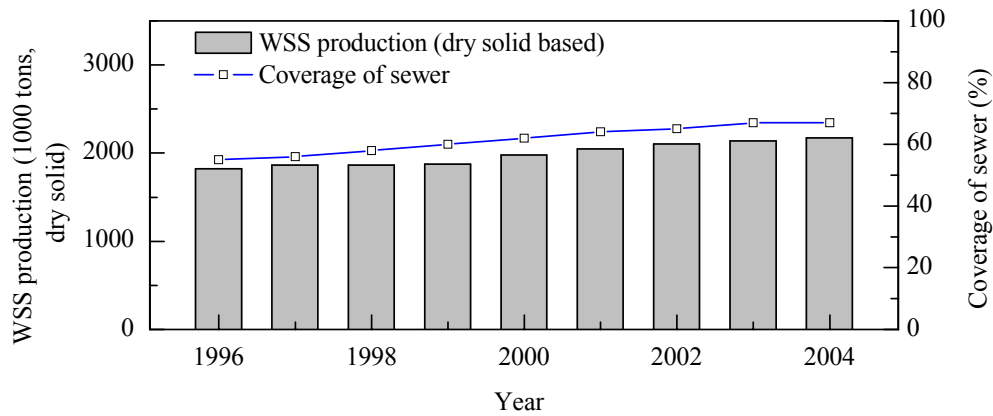


Figure 2.5 Sewer coverage and wasted sewage sludge production in Japan (Japan Ministry of the Environment, 2012b)

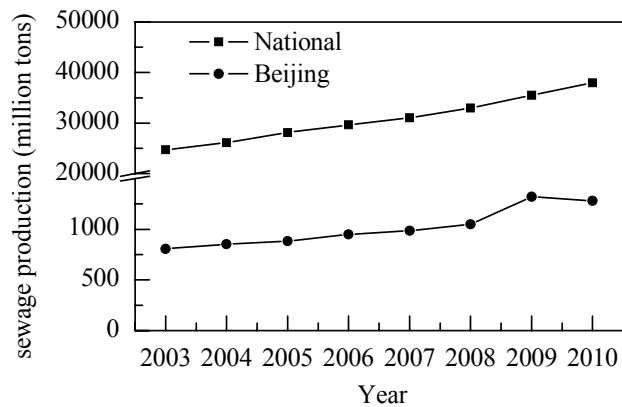


Figure 2.6 Sewage production in China: National and Beijing (National Bureau of Statistic of China, 2012)

2.1.4 Strategies for municipal biowaste disposal

Parts of MBW like papers can be recycled or reused, while other MBW needs further disposal. OFMSW like RKW and FVW are usually mixed and disposed together with other MSW, while WSS is usually separately disposed.

Landfill, incineration and composting are now the main strategies for MSW disposal worldwide. Figure 2.7 gives the proportions of MSW disposed by each method in different areas. In United States, Europe and China, landfill is most widely used. But for Japan, incineration is the most popular disposal method, probably due to its scarcity of landfill site. For the case in China, only 66.8% of the collected MSW were disposed in 2008 due to scarcity of related facilities (National Bureau of Statistic of China, 2012). The untreated MSW were dumped in open areas and caused serious environmental concern.

To be mentioned, disposal of RKW and FVW in China was out of control to some extent. Most of them were directly used as livestock feed before, without control of the components and bacteria. By now, RKW has been forbidden to use as feed of livestock directly. Due to the special dietary habit in China, the RKW usually has high contents of oil which can be recycled as fuel oil. But most of this recycled oil is actually sold as edible oil illegally. According to these situations, more efforts are needed for the correct collection and disposal of RKW and FVW.

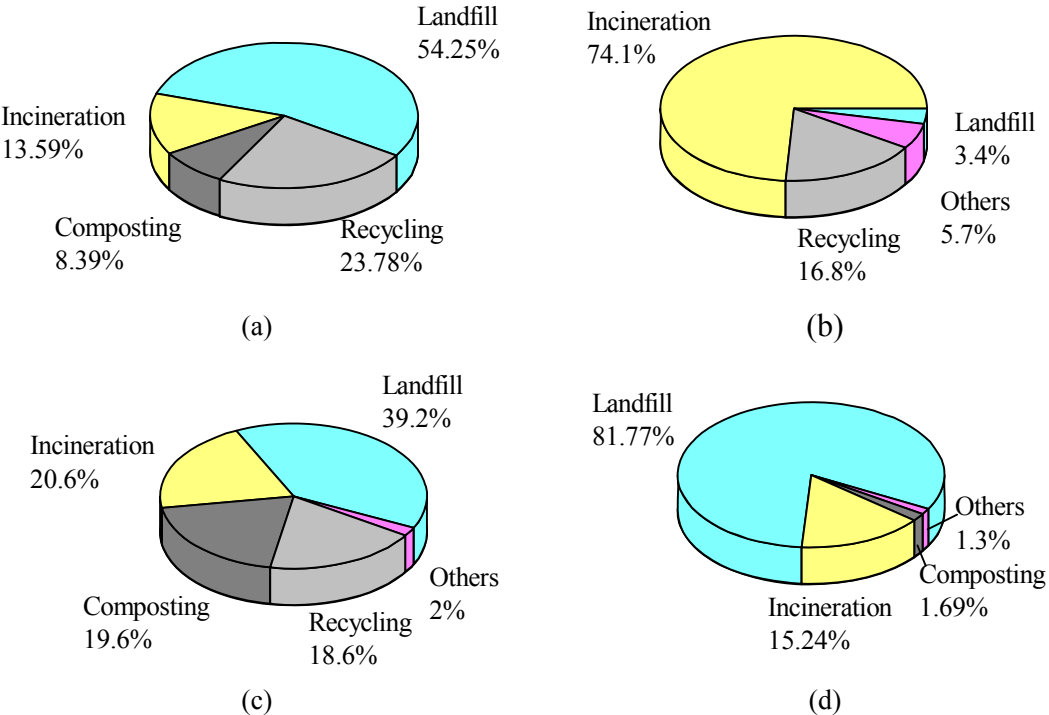


Figure 2.7 Disposal of MSW in different areas: (a) United states, 2005 (OECD, 2008); (b) Japan, 2005 (OECD, 2008); (c) EU-15, 2003-2005 (OECD, 2008); (d) China, 2008 (National Bureau of Statistic of China, 2012)

WSS is usually separately collected and disposed. In Japan, about 70% of WSS were treated by incineration while the left 30% by landfill and other methods (MLIT, 2012). However, it was reported that only 10% of the WSS in China was disposed of by incineration. Another 6% was disposed of by landfill and 1% by composting. The remaining 83%, dumped without any treatment, presents a considerable environmental risk (China water net, 2012).

MBW contains high content of water and biodegradable organic matters. Due to these special properties, it causes secondary pollution and other problems when disposed in the above traditional strategies (Hartmann and Ahring, 2006).

MBW causes emission of greenhouse gas (GHG) and high-strength leachate during landfill. USEPA estimated that 5.6 million t of methane was produced from landfill in the United States in 2009, which was 17% of the total methane emission from human-related sources in United States (USEPA, 2012b). EU has estimated that GHG emission in EU-25 can be over 100 million t CO₂-equivalents in 2020 (European Environmental Agency, 2012). To release these problems, several measures have been taken. For example, EU has published the Council Directive 99/31/EC. It is required that MBW going to landfills should be reduced to 50% of that in 1999 by 2009 and 35% by 2016 (European Council, 1999). Reduction of the MBW before landfill can also release the problem of land scarcity, which is especially appreciated in Japan and some megacities, where land resources are very rare.

For the incineration processes, MBW can also be a bad factor. Due to the high water content in MBW (usually 60%-90%), extra fuels are needed to avoid unsteady burning and formation of dioxins when MBW are mixed with other MSWs (Kanters and Louw, 1994). Hartmann and Ahring (2006) suggested that only the non-degradable MSW should be disposed by incineration while MBW should be treated firstly by anaerobic digestion.

Aerobic composting represents an energy consuming process (around 30–35 kWh is consumed per t of waste input). It occupied large areas and causes odor problems (Braber, 1995). Moreover, composting usually occurs in open areas, which causes considerable GHG emission (Edelmann et al., 2000). It was accounted that 0.08 million t of methane were produced from composting in the United States in 2009 (USEPA, 2012b).

Due to the disadvantage of MBW disposal by the above methods, improved strategies have been brought out for its disposal. Among them, anaerobic digestion (AD) was suggested for the treatment of MBW before final disposal by landfill, incineration or composting. As MBW has high content of biodegradable organics, it can be a bio-energy resource and the high water content makes this bio-energy feasible to be recovered by AD processes (De Baere, 2000 and 2006; DiStefano and Belenky, 2009; El Hanandeh and El-Zein, 2010).

2.2 Anaerobic digestion: principle and application

2.2.1 Basic principles of anaerobic digestion

Anaerobic digestion is a series of processes in which microorganisms break down biodegradable material in the absence of oxygen. It has some advantages over other biological processes, such as high organic loading rate and low sludge production rate. Most importantly, AD processes can release the bio-energy in the waste while other biological processes are usually energy-cost. AD has been utilized by human to process food and beverage since ancient times. Scientific explanation and utilization of AD processes began since the 17th century. In 1930s, anaerobic bacteria were discovered as microorganisms promote the AD process. Buswell and Neave (1930) firstly divided the AD process into two sections. After that, many models have been developed to explain the AD processes. In 2002, International Water Association (IWA) published a mathematical model named Anaerobic Digestion Model No.1 (ADM1) (Batstone et al., 2002). This model is widely accepted as a summary of the former researches and gives the most complete description of the AD processes. However, the AD processes are so complex that this model still has no numerical solution if not simplified.

In ADM1 model, AD processes are presented as the following five steps: disintegration, hydrolysis, acidogenesis, acetogenesis, and methanogenesis. Figure 2.8 gives the basic mass flow of the AD processes.

In the first step of disintegration, complex particulates were degraded into particulate organics, including carbohydrates, proteins, lipids. This process is mainly catalyzed by extracellular enzyme, but not the microorganisms. Carbohydrates, proteins and lipids are then hydrolyzed by extra- and intracellular enzymes. The intermediate products are acidified into short-chain fatty acids in the step of acidogenesis, and furtherly into acetate in the step of acetogenesis. Methane is then formed in the final step of methanogenesis. A part of the methane is produced from acetate, facilitated by acetoclastic methanogens. Other part of the methane is built up from hydrogen and carbon dioxide, which is processed by hydrogenotrophic methanogens.

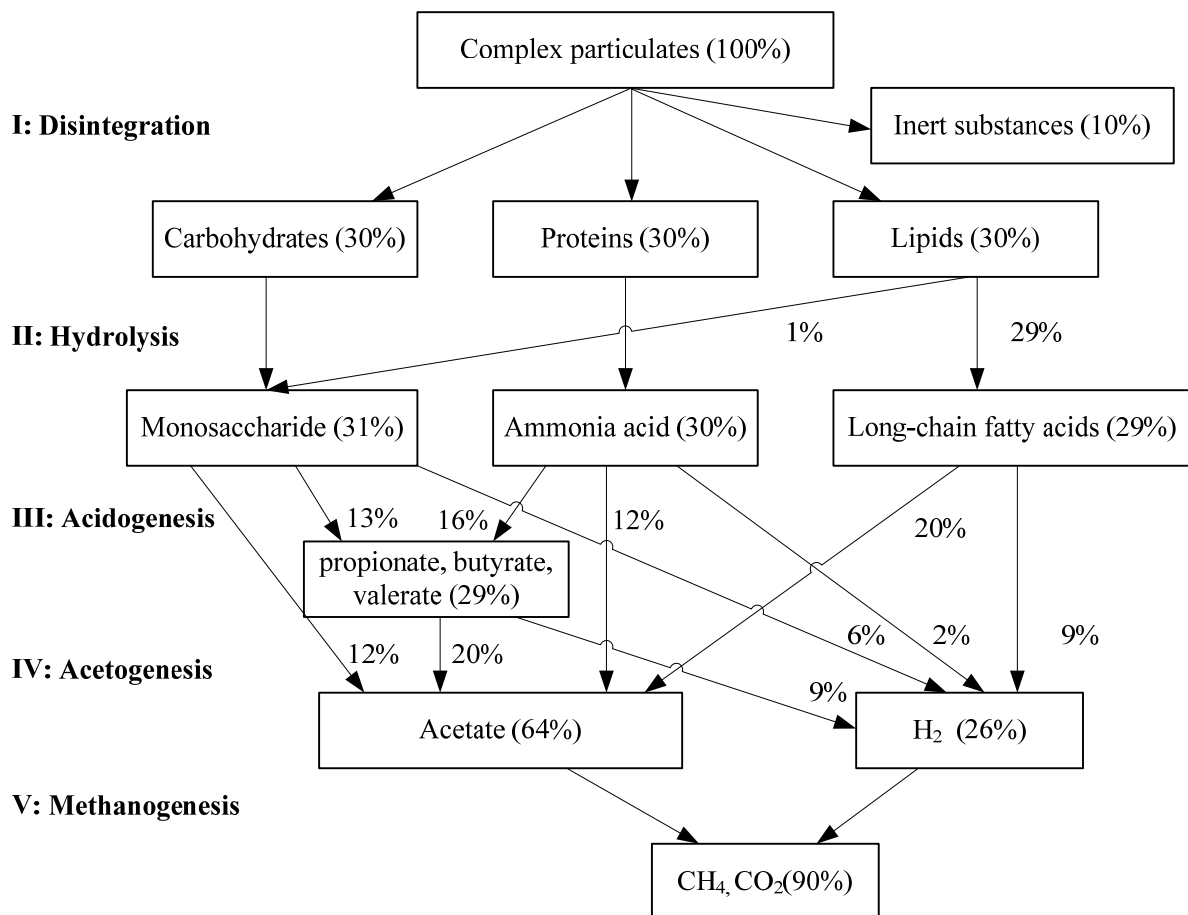


Figure 2.8 Anaerobic digestion processes (Batstone et al., 2002)

2.2.2 Anaerobic digestion of municipal biowaste: application

AD has been widely used in the wastewater treatment processes, while application of AD processes on MBW treatment is not so popular. These years, its application on MBW treatment is attracting more and more attention, due to the limitation in other disposal strategies, as discussed in Section 2.1.2. Anaerobic digestion (AD) of MBW has been employed in Western Europe since the 1980s. Its economic feasibility was also proved, considering environment benefit and increasing fees of traditional waste treatment technologies (Edelmann et al., 2000; DiStefano and Belenky, 2009; El Hanandeh and El-Zein A, 2010).

Application of AD technology on MBW treatment is most popular in Europe, which was developed since 1980s and early 1990s. De Baere (2006) reported that 124 full-scale plants had been built by 2006, with a total digestion capacity of about 4 million t per year. The construction rate of these facilities was also increasing. An average annual increase of

10.4 plants was found from 2001 to 2005, with a capacity growth rate of 428,000 t per year. As a comparison, the annual increase was only 2.6 plants with a capacity of 33,000 t per year from 1991 to 1995.

In Japan, anaerobic digestion was widely used for WSS and also agricultural waste, while the commercial application on other MBW is limited.

Four anaerobic digester systems was built using the AD technology Wassa (developed by Finland) in Tokyo, Ikoma, Shimoina, and Jouetsu, which together treated over 20,000 t of dewatered WSS and other MBW per year. A plant treating 20,000 t per year was also built by another European company (Kompogas, Switzerland) in Kyoto in 2004. In Hita city, a project named 'biomass town' has been brought up since 2005. MBW are digested in an AD system with a capacity of 80 t/d to produce electricity and heat (MAFF, 2005). In Suzu city, similar project was also carried out since 2005 and it was estimated treatment capacity of the AD system can reach up to 51 t/d in 2012 (MLITT, 2012).

In China, AD technology is most widely applied for the treatment of manure in rural area and WSS in urban area. Several large-scale plants have been built treating other MBW these years. Construction of the first plant began in Shanghai in 2005, which has a treatment capacity of 280,000 t per year. Other two plants were built in Beijing and Guangzhou, with capacity of about 240,000 t and 350,000 t per year. The one in Beijing treats 650 t of waste every day, including 150 t of selected OFMSW, 300 t of MSW and 200 t of food waste.

2.2.3 Anaerobic digestion of municipal biowaste: economic, environmental and energetic feasibility

Economic, environmental and energetic feasibility of AD processes treating MBW has also been proved.

The treatment capacities for AD in Europe are, however, still 20 times less than for aerobic composting, probably because AD was not considered to be a fully proven technology until around 1995. It was also more expensive, so many municipalities chose less risk and less investment (De Baere, 2000). With an increasing internalization of external costs, however, the positive environmental effects are considered within economical constraints. Economical feasibility can alter radically with energy prices, levies on waste disposal and land prices

(Braber, 1995). In a life-cycle assessment, Edelmann et al. (2000) showed that AD was advantageous compared to composting, incineration or to a combination of digestion and composting, mainly because of a better energy balance. El Hanandeh et al. (2010) used a multi-criteria decision-making tool named ELECTRE-SS to select a management strategy for biodegradable fraction in the municipal solid waste of Sydney, and found alternatives based on AD ranked highest.

2.3 Anaerobic digestion of municipal biowaste: limitation and solutions

Although AD technology has become well established and accepted as a treatment method of MBW, its application is still limited. There were many reasons that blocked the development of AD on MBW treatment, including technical and economic aspects (De Baere, 2000; Stroot et al., 2001). Among them, the limited treatment efficiency is one of the most concerned. It leads to unstable operation of the digesters and decreased treatment capacity.

Digestion efficiency is mainly limited by slow hydrolysis rate of solids and growth rate of anaerobic microorganisms.

2.3.1 Limitation from solid hydrolysis and its solution

Solid disintegration and hydrolysis is usually thought as the limiting step for digestion of solid organics (Eastman and Ferguson, 1981). Compared with wastewater, MBW has a higher solid concentration of about 30% TS (Hartmann and Ahring, 2006). As a result, the limitation from solid hydrolysis is more serious in anaerobic digestion of MBW than wastewater.

MBW has a complex composition. Among its different components, fibers from yard waste, FVW, and papers have very low hydrolysis rates. The presence of lignin and hemicellulose makes the access of cellulase enzymes to cellulose difficult, thus reducing the efficiency of the hydrolysis (McMillan, 1994). WSS may also be a limiting component as most of the organics in WSS present in the form of microbial cells. Destroying of the cell wall and membrane takes a long time and thus becomes the limiting step in the digestion of WSS (Müller, 2000; Nah et al., 2000).

Different methods were developed to increase the hydrolysis rate of MBW (Gunaseelan, 1997; Müller, 2000; Mata-Alvarez et al., 2000; Nah et al., 2000), including physical, chemical, biological, and combined pre-treatment before anaerobic digestion.

1) Physical pre-treatment

Physical pre-treatments include mechanical crashing, ultrasonic irradiation, thermal hydrolysis, and so on.

Mechanical crashing can diminish the particle sizes and destroy the structure of cells and particles. Digestion efficiency and biogas production can then be improved. An increase of 17%-25% in biogas production was found when mechanical crashing was applied as pre-treatment (Hamzawi et al., 1999; Angelidaki and Ahring, 1994).

Ultrasonic irradiation enhances MBW digestibility by disrupting structure of cells. Intracellular organics are then released and digestibility is improved (Tiehm et al., 1997; Hogan, 2004; Appels, 2008). It was reported that biogas production from WSS can be improved by 10%-48% (Appels, 2008).

Thermal pre-treatment can help to release organics inside the cells and consequently improve digestion performance (Phothilangka, 2008). Thermal pre-treatment has been operated under a temperature range of 60-270 °C, with the optimal range of 160-180°C and treatment times from 30 to 60 minutes (Weemaes and Verstraete, 1998). Haug et al. (1978) , Li et al. (1992), and Sawayama et al. (1997) found the most suitable temperatures were 170-175 °C, at a pressure of around 10 bar . A 40%-100% increase of biogas production was found for WSS after thermal pre-treatment (Zhang, 2010).

Physical pre-treatments can be simply operated and have short treatment times. However, they usually need high energy input and cost.

2) Chemical pre-treatment

In chemical pre-treatments, different chemicals are added to improve hydrolysis, including acids, alkaline, ozone, and so on.

Sun and Cheng (2005) used diluted sulfuric acid for the pre-treatment of straws. With a sulfuric acid concentration higher than 1.2% and treatment time over 60 minutes, 50% - 66% of hemicellulose was hydrolyzed into monosaccharide, which can be much easier to be

digested. However, this method is limited due to large consumption of acids for pre-treatment and alkali for pH adjustment in further AD treatment.

Addition of alkali is more widely used as pre-treatment before AD compared with addition of acids, as the suitable pH values for AD processes are 6.8-7.2 (McCarty and McKinney, 1961). Pavlostathis and Gossett (1985) found alkali addition resulted in an over 100% increase in digestibility of wheat straw.

Müller (2000) compared ozonation with other pre-treatments like heating, mechanical crashing, and so on. He found ozonation had the highest degree of disintegration of WSS. 50% of total COD can be disintegrated into sCOD under an energy consumption of 5 MJ/kg SS, while mechanical crashing by stirred ball mill can only disintegrate 35% of total COD under the same energy input. A 16%-145% increase of biogas production was also found for WSS after ozonation by Zhang (2010).

Despite of its benefit on digestibility, chemical pre-treatment may cause corrosion of facilities due to agent addition. It also increases operation complexity and cost.

3) Biological pre-treatment

Biological pre-treatments include aerobic composting, anaerobic pre-treatment, and so on.

Many researchers have studied the effect of aerobic composting on digestion efficiency. Hasegawa et al. (2000) found biogas production from organic sludge was improved by 50% after thermophilic aerobic composting. Fdez-Güelfo et al (2011) also found an increase of 190% in ultimate biogas production. Charles et al. (2009) found aerobic pre-treatment can help with the start-up of anaerobic digesters.

In some cases, hydrolytic bacteria were directly added to improve digestion efficiency. Del Borghi (1999) found biogas production increased 30%-100% when 1.0 g/L of hydrolytic bacteria was added for pre-treatment.

Anaerobic pre-treatment mainly occurred as the acidification phase in two-phase digesters. In this type of digesters, hydrolysis can then be optimized in the first phase under a lower pH (De Baere, 2006).

2.3.2 Limitation from microbial growth and its solution

Another factor that limited the development of AD processes is the slow growth rate of anaerobic bacteria.

As described in Figure 2.7, anaerobic digestion is comprised by complex processes and affected by different groups of microorganisms. However, methanogens, which produce methane, are the most important groups. These bacteria have very slow growth rate, especially acetoclastic methanogens (Rittmann and McCarty, 2001). As a result, long retention times are needed for the anaerobic bacteria. This means the volume of digesters should be enlarged or the treatment time should be prolonged. Both of them bring disadvantages to the application of AD technologies.

To solve this problem, several solutions have been brought up.

One typical example is the development of two-phase digesters instead of single-phase reactors. As discussed before, hydrolysis/acidification and methanisation are processed in two different reactors in a two-phase digester. The two reactors then can be independently conditioned for better growth of different microorganisms. Several commercial two-phases AD processes were developed, like Pacques process in Netherlands, BTA processes in Germany and Canada, Biocomp and Biopercolate processes in Germany. However, the practical use of two-phase digesters was limited due to complex design and operation of the two reactors. De Baere (2006) suggested that more than 87% of the digestion capacity in Europe were still provided by single-phase reactors.

Co-digestion of different feedstock was also taken into consideration in many references. Feedstock properties like biodegradability, C/N ratio, moisture and so on will be changed, diverting to higher microbial growth rates and better digestion performance compared with single feedstock (Stroot et al, 2001; Hartmann and Ahring, 2005; Bouallagui et al., 2009; Holm-Nielsen et al., 2009; Li et al, 2010). Co-digestion of different feed also made feedstock supply easier, especially for those large-scale plants, where large quantities of MBW were needed. Mata-Alvarez et al (2000) have summarized several cases of co-digestion of MBW, in both research work and practical use.

Some other strategies are also applied to improve the growth of typical bacteria. Selection, incubation and immobilization of dominant bacteria was also thought noteworthy (Lopes et al., 2004; Forster-Carneiro et al., 2007).

2.4 Application of thermal hydrolysis as pre-treatment before anaerobic digestion

Among the above solutions for hydrolysis limitation, thermal hydrolysis is one of most studied solutions. It was firstly proved beneficial for digestibility and dewaterability of sewage sludge in 1971 (Fisher and Swanwick, 1971). After that, many researches were done for the application of thermal hydrolysis. Commercial technologies have also been developed based on thermal hydrolysis, known as Cambi™ by a Norwegian company (1995) and BioTHELYS™ by a French company (2006).

The Cambi process treats the organic matter at 165-170°C, dissolving it into an easily digestible feed for biogas production by anaerobic digestion, while destroying any harmful organisms (pathogens). It has already been operated in 17 commercial sewage treatment plants (Cambi™, 2012). Organic removal rates of 50%-65% can be achieved and dewaterability can be improved by 50%-100%. Net energy production can also increase 20% with this process.

BioTHELYS™ technology has been used in four commercial projects. With treatment temperatures ranged from 150°C to 170°C and treatment times of 30-60 minutes, biogas productions were improved by 60%-70% (Gong et al., 2010).

Most of these researches were limited to treatment of WSS. However, it was found that types of the feeding materials significantly affected the performance of thermal hydrolysis (Qiao et al., 2012). As a result, it's still meaningful to evaluate the effect of thermal hydrolysis on other MBWs.

2.5 Anaerobic sequencing batch reactor (ASBR)

ASBR was firstly brought up by Dague et al. (1992). Four phases are included in one treatment cycle of the ASBR, as shown in Figure 2.9.

Continuously stirred tank reactor (CSTR) is now most widely used in AD processes. With the settling phase included, ASBR may outperform CSTR in many aspects:

1) Better effluent quality can be achieved. As most suspended solids are settled, much less solids are contained in effluent.

2) Higher digestion efficiency. As solid retention time (SRT) is derived from hydraulic retention time (HRT), digester efficiency can be improved by enhanced solid hydrolysis. Digester efficiency is also improved as more microorganisms are kept inside the digester.

3) Higher stability. As solids are kept inside the digester, solid concentration can reach to a high level in ASBR. This in turn improves stability of the digesters.

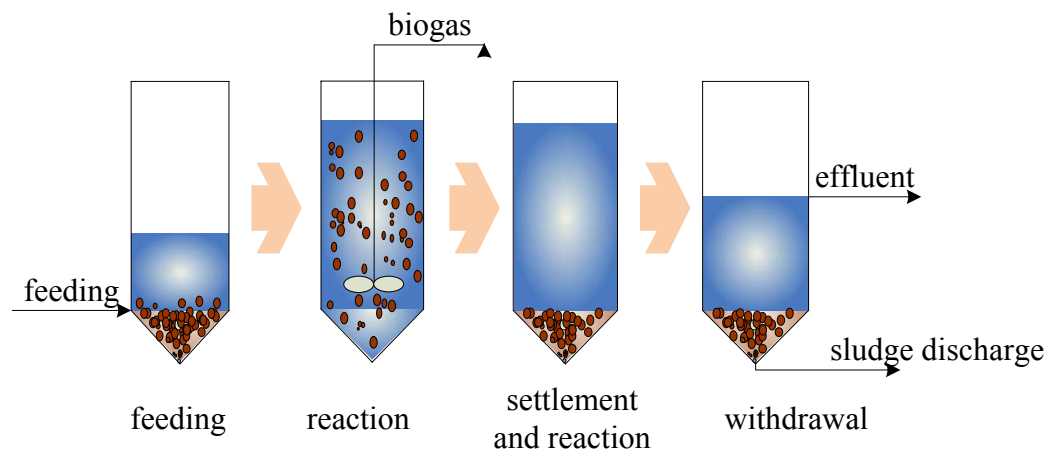


Figure 2.9 Scheme of ASBR

As ASBR can be easily switched from CSTR without structure variation of the digesters, cost can be saved for construction. Operation of ASBR is also simpler than some other digesters like up-flow anaerobic sludge blanket (UASB).

Zaiat and his fellows did a series of work on ASBR, considering the effect of feeding strategy (Ratusznei et al., 2003), agitation rate (Rodrigues et al., 2003; Pinho et al., 2004; Cubas et al., 2004; Farias de Novaes et al., 2010), organic loading rate (Siman et al., 2004; Mockaitis et al., 2006; Damasceno et al., 2007), and addition of bio-film carriers (Siman et al., 2004; Sarti et al., 2007). They found ASBR can be more stable and effective than CSTR treating wastewater, with COD removal rates over 90% under most occasions.

The major disadvantage of ASBR could be unstable settling performance at high organic loading rates and/or high sludge concentrations. Under this condition, poor solid-liquid separation leads to lost of microorganisms, increase of effluent concentrations, and finally

unstable operation of the digesters. This disadvantage has blocked the application of ASBR for MBW digestion, as MBW usually has a much higher solid concentration than wastewater.

Hur (1999) and Luo (2009) reported success operation of ASBR treating high-solids-content MBW. However, their researches were only in lab-scale and solid-liquid separation was fulfilled by floatation in Hur's case. One of the research groups in this study, Wang and his fellows, found thermal hydrolysis can not only improve digestibility but also the settling performance of MBW and its digestate (Wang Z.J. and Wang W., 2005; Wang et al., 2009). ASBR was used as digester after thermal pre-treatment in their studies. However, these studies were also mainly limited to lab-scale and pilot validation is needed in following studies.

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Chapter 3 Effect of thermal hydrolysis on properties of municipal biowaste

3.1 Introduction

Anaerobic digestion of municipal biowaste (MBW) is usually rate-limited by solid hydrolysis. Thermal hydrolysis has been proved one of the useful pre-treatment methods to improve the solid hydrolysis rate (Fisher and Swanwick, 1971; Haug et al., 1978; Mata-Alvarez et al., 2000). It's thought that thermal hydrolysis can break cell structure to release organics inside the cells and consequently improve digestion performance (Phothilangka et al., 2008).

Several commercial technologies have also been developed based on these studies, using thermal hydrolysis as pre-treatment before anaerobic digestion. Successful examples include Cambi™ and BioTHELYS™. However, most of the applications are located in Europe and mainly focus on the treatment of sewage sludge. Studies and applications in other areas and other MBW are lack.

Meanwhile, co-digestion of sewage sludge with other types of MBW is becoming popular these years, as co-digestion can improve biogas output and stability of the anaerobic digesters (Hartmann and Ahring, 2006; De Baere, 2006; Bouallagui et al., 2009). Types of MBW may significantly affect the performance of thermal hydrolysis (Qiao et al., 2012).

Restaurant kitchen waste (RKW) and fruit-vegetable waste (FVW), which have large productions and are easy to be separately collected, were mixed with sewage sludge to improve anaerobic digestion efficiency in this study. Thermal hydrolysis was used as pre-treatment. As no similar studies were found on these mixed MBW in China, it's meaningful to find the effect of thermal hydrolysis on them.

Former studies in the same group have found the effect of thermal hydrolysis on these mixed MBW in lab scale (Wang et al., 2009; Liu et al., 2012). However, there are some differences between lab- and pilot- scale thermal hydrolysis processes. Generally, electric power is used as heat source in lab scale thermal hydrolysis whereas steam is used in pilot or industrial plants (Chauzy et al., 2005; Mottet et al., 2009). Use of steam as heat source can avoid the problems of local overheating when the facilities are scaling up. The cost will also

be reduced as cheaper fuels can be used to produce steam rather than electricity (Kepp et al., 2000). In this study, experiments were carried out in a pilot plant. Steam was used as heat source while electricity was used in former lab studies. As a result, it's also important to compare these two types of thermal hydrolysis processes.

Based on the above information, the objective of this chapter was to find out the effect of pilot-scale thermal hydrolysis on properties of the mixed MBW. Results were compared with other studies, especially with the former lab studies.

3.2 Materials and methods

3.2.1 Municipal biowaste

As mentioned in Chapter 2, three different types of municipal biowaste were selected, RKW from a University canteen, FVW from a local food market, and wasted sewage sludge (WSS) from a sewage treatment plant in Beijing. Properties of these wastes were listed in one of the referenced papers (Liu et al., 2012) and were cited in Table 3.1.

Table 3.1 Properties of targeted MBW (average values)

	RKW	FVW	WSS
Water content (%)	80.01	89.26	84.26
TS (g/kg)	199.93	107.41	157.42
VS (g/kg)	180.43	100.42	114.06
SS (g/kg)	87.71	60.78	149.48
VSS (g/kg)	83.85	55.11	108.28
Crude fibre (g/kg)	10.57	15.37	14.18
Crude protein (g/kg)	30.24	13.23	68.68
Crude fat (g/kg)	43.63	3.10	20.52
C (%-dry based)	48.69	46.91	38.65
H (%-dry based)	7.25	6.37	6.02
O (%-dry based)	32.62	40.71	26.30
N (%-dry based)	2.82	2.16	5.57
C/N ratio	17.3	21.7	7.0

Each of these wastes was collected once or twice weekly, accounting for about 500 kg each time. After that, they were shredded separately with a hammer crusher (A300, BestPower, China) to an average size of less than 3.0 mm and mixed in a weight ratio of 1/1/1. This kind of mixed waste was named rMBW. This rMBW was then treated by thermal hydrolysis, or directly digested.

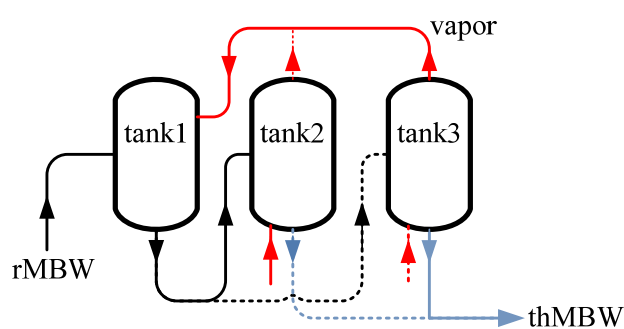
3.2.2 Pilot-scale thermal hydrolysis facility

Pilot-scale thermal hydrolysis was carried out for one day every 1 or 2 weeks. The process was carried out under a temperature range of 170-175°C for 60 minutes. Water steam in a pressure of 0.5 MPa was used as heat source, while the highest pressure in the hydrolysis tank can reach to about 1.2 MPa. Temperature was controlled by an automatic system.

The facility is as shown in Figure 3.1. Three tanks are included, with an effective volume of 80 L respectively. Tank 1 was used as pre-heating. Tank 2 and 3 were used for thermal hydrolysis in turn. When thermally hydrolyzed MBW (thMBW) were released from tank 2 or 3, pressure in these tanks crashed down immediately and thus flash evaporation occurred. The vapor was released back to tank 1 for pre-heating the biowaste to about 95 °C.



(a)



(b)

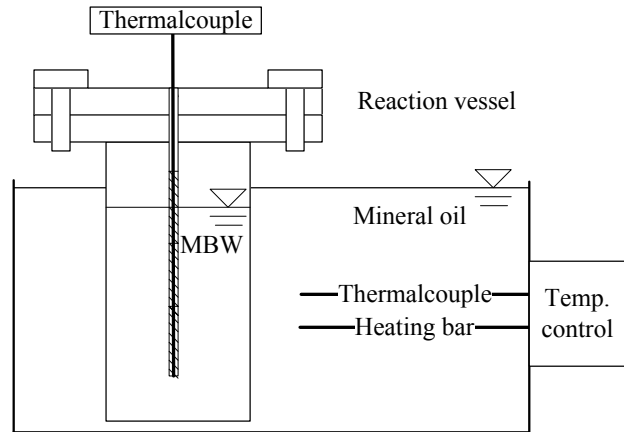
Figure 3.1 Pilot-scale thermal hydrolysis facility: (a) photo; (b) sketch (in the case that tank 2 was used for thermal hydrolysis, while tank 3 for flashing; dot line means not in use under this case)

3.2.3 Lab-scale thermal hydrolysis facility

To clarify the difference between pilot- and lab-scale thermal hydrolysis, the equipment used for thermal hydrolysis in former lab-scale studies was also introduced. It's as shown in Figure 3.2. MBW was heated by oil bath, whose heat was provided by electricity.



(a)



(b)

Figure 3.2 Lab-scale thermal hydrolysis facility: (a) photo; (b) sketch (one of the vessels)

From Figure 3.1, it can be found, MBW got contact with the steam in pilot-scale facility. Accordingly, MBW got diluted by the water steam. However, reaction happened in a closed vessel in lab facilities, as shown in Figure 3.2.

3.2.4 Sampling and analysis

1) Sampling

Samples of rMBW and thMBW were collected once or twice a month, lasting about 6 months. Both of them were measured for their biochemical and physical properties. Biochemical indicators include: solid concentrations, soluble chemical oxygen demand (sCOD), pH, VFA, alkalinity, ammonia, and biochemical methane potential (BMP). Physical indicators include: particle size distribution, viscosity, dewaterability, and settling performance. Result of the above parameters was an average of more than 8 times determination except for BMP and particle size distribution.

2) BMP test

BMP test was applied for the digestibility determination of rMBW and thMBW, following the instructions given by IWA (Angelidaki et al., 2009). Sludge from one of the pilot digester was used as inoculums and pre-incubated to remove the residual biodegradable organics in it. The pre-incubation was carried out in 35°C water bath for 2 d.

The test was then carried out in 100 ml flasks. Each flask was added with 75 ml inoculums and a certain amount of rMBW or thMBW to meet a $VS_{\text{substrate}}/VSS_{\text{inoculum}}$ ratio of

1/4. Blank assay was also carried out by no addition of any feed and all the samples had three parallels. Flasks were then added with water to meet a total volume of 90 ml. All flasks were purged with nitrogen gas before sealing to create anaerobic conditions. All flasks were kept in water bath under temperature of 35 ± 1 °C and fully shaken every 12 hours by hand.

Production of biogas and its methane content was detected until no biogas was produced. Biogas was collected by graduated cylinders. Methane content was monitored by a gas chromatography system (GC-2010, Shimadzu, Japan) with a $30 \text{ m} \times 0.53 \text{ mm}$ I.D. capillary column (RT-Qplot, Restek, USA) and a thermal conductivity detector.

3) Analysis of other biochemical indicators

Solid concentrations can be concluded from the results of 4S, which means total solids (TS), volatile solids (VS), Suspended solids (SS) and volatile suspended solids (VSS). TS stands for contents of all the substances expect for water. Its composition can be described as Equation 3-1 and 3-2 (Wang et al., 2009).

$$TS = SS + DS = (VSS + FSS) + (VDS + FDS) \quad (3-1)$$

$$TS = VS + FS = (VSS + VDS) + (FSS + FDS) \quad (3-2)$$

FS: fixed solid; FSS: fixed suspended solid; FDS: fixed dissolved solid; DS: dissolved solid; VDS: volatile dissolved solid.

Value of pH was determined by a lab-use pH meter (FE20, Mettler-Toledo, Switzerland). Alkalinity and ammonia was detected by titration. The rMBW or thMBW was firstly distilled before ammonia detection. The sCOD was detected according to standard methods (APHA-AWWA-WEF, 2005). Volatile fatty acids (VFAs), including acetic, propionic, iso-butyric, butyric, iso-valeric, and valeric acids, were determined by a gas chromatography system (GC-2010, Shimadzu, Japan) equipped with a capillary column (RTX-1, $15 \text{ m} \times 0.53 \text{ mm} \times 0.50 \text{ mm}$, Shimadzu, Japan) and a flame ionization detector. Samples used for sCOD and VFA determination were firstly filtered through $0.45 \mu\text{m}$ filters and acidified with same amount of formic acid.

4) Particle size distribution

Particle size distribution of rMBW, thMBW was analyzed by sieving method. The method was established based on a publication from Laguna et al. (1999), with a little change. Six stainless sieves with diameters of 10 cm were used, with mesh sizes of 900, 450, 200, 125, 98 and 74 μ m. About 10 g of MBW was used for analysis. Each sieve was carefully washed using about 100 ml of ionized water during the sieving processes. After that, the sieved solids were back-washed off the sieves using about 400 ml of ionized water. The back-wash water were then collected and filtered by a 0.45 μ m filter for solid weighting.

5) Analysis of other physical indicators

Viscosity was measured by a rotational viscometer (NDJ-1, Shanghai Jingke, China). Dewaterability was represented by both capillary suction time (CST) and dryness of dewatered waste. CST was determined by analyzer (Triton 304M, Triton Electronics, England). The waste was also dewatered by a frame filter pressing system (pore size of filter: 20 μ m) under 1 MPa of pressure for 30 min. Dewatered cake was detected for their dryness, which was also used as a representative of dewaterability. Settling performance was represented by settling velocity, which is detected in a 500 ml vessel.

3.3 Effect of thermal hydrolysis on biochemical properties of MBW

3.3.1 Solids and sCOD

TS, SS, VS, and VSS were measured directly. The results were shown in Table 3.1. As thMBW was diluted by water vapour during thermal hydrolysis, solids contents should be revised for better comparison. By supposing no TS were lost during hydrolysis, a dilution fold of 1.6 was calculated. Values of VS, SS, and VSS were then multiplied by 1.6 to get rid of the effect of dilution. The revised results were also listed in Table 3.2.

After thermal hydrolysis, VS decreased a little. This may be due to evaporation of some substances during thermal hydrolysis. However, this decrease was not so significant and was ignored in the following discussion. Especially, three indicators were used for discussion, including VSS hydrolysis ratio, VDS/VS ratio, and sCOD increase. They can represent the degree of hydrolysis from suspended organics into soluble organics.

Table 3.2 Solids and sCOD in rMBW and thMBW

	rMBW	thMBW	thMBW (revised)
TS (g/kg)	109.3 ± 14.3	68.7 ± 14.3	109.3 ± 22.7
VS (g/kg)	87.0 ± 11.7	52.2 ± 11.0	83.0 ± 17.4
SS (g/kg)	86.2 ± 13.0	37.9 ± 14.2	60.2 ± 22.5
VSS (g/kg)	70.9 ± 11.0	27.5 ± 8.8	43.7 ± 14.1
VDS/VS	0.19 ± 0.03		0.48 ± 0.10
sCOD (g/l)	39.6 ± 8.0	38.7 ± 7.0	61.6 ± 11.1

VSS, which represents the content of organic suspended substances, can be calculated as following:

$$\text{VSS hydrolysis ratio} = \frac{\text{VSS in rMBW} - \text{VSS in thMBW(revised)}}{\text{VSS in rMBW}} \times 100\% \quad (3-3)$$

As VSS were hydrolyzed into VDS, VDS/VS ratio can also be used to represent the degree of hydrolysis as well. This parameter is not related to the dilution effect, so it's simpler for the evaluation of hydrolysis in this study.

Moreover, sCOD, which is more widely used to represent the concentration of dissolved organics, was also discussed for its increase from solid hydrolysis.

It was found VSS hydrolysis ratio was 38.3%. It led to a VDS/VS ratio increase from 0.19 to 0.48 and a sCOD increase of 56.0%. These results were comparable with other studies (Haug et al., 1978; Kim et al., 2005; Wilson and Novak, 2009). In lab studies, it was found the VSS hydrolysis ratio was 38.9%, 38.4% and 27.5% for RKW, FVW, and WSS. The VDS/VS ratio was increased by 0.15, 0.17, and 0.21 for FVW, RKW, and WSS (Liu et al., 2012). These increases were lower than that obtained in this study (0.29).

It can be found thermal hydrolysis was a little more efficient in the pilot-scale study than in the lab studies. This phenomenon was explainable. In pilot facilities, steam was used as heat source, which can directly contact with MBW. While in lab studies, oil bath was used for heating and thus efficient heat transfer was needed. As MBW was not a homogeneous

medium, unequal heating might happen locally during thermal hydrolysis and the reaction might be weakened consequently.

3.3.2 Inhibitors and related parameters: VFA, ammonia, alkalinity and pH

VFA and ammonia are common inhibitors during anaerobic digestion. Meanwhile, both of them are related to alkalinity and pH as they are also important buffers in the digestate. Usually, VFA/alkalinity value of less than 0.3 and ammonia concentration of lower than 1.5-3 g N/l is suggested for stable operation of anaerobic digesters. An optimal pH range of 6.8-7.2 is also suggested (McCarty and McKinney, 1961).

Feeding materials with inappropriate values of these parameters may cause inhibition and affect the performance anaerobic digesters. As thermal hydrolysis was applied as pre-treatment before anaerobic digestion, it's important to look into the variation of these parameters during thermal hydrolysis.

Table 3.3 listed variation of these parameters after thermal hydrolysis.

Table 3.3 VFA , alkalinity, and pH in rMBW and thMBW

	rMBW	thMBW	thMBW (revised)*
VFA as COD (mg/l)	10092 ± 1006	9035 ± 1787	14365 ± 2841
Ammonia (mg/l)	1518 ± 144	731 ± 57	1163 ± 90
Alkalinity (mg/l)	4629 ± 590	3704 ± 868	5889 ± 1380
pH	4.41 ± 0.21	4.77 ± 0.44	4.57 ± 0.23

*supposing that thMBW was ideal solution, ionization re-equilibrium ignored

The data were also revised for dilution. During calculation, the thMBW was thought as an ideal solution and effect of ionization re-equilibrium was ignored. As pH variation was not so significant, this calculation was still useful to some extent.

Effect of thermal hydrolysis on the change of VFA and ammonia concentrations was not significant. Ammonia concentration decreased a little, probably resulted from measuring error and evaporation into the pre-heating tank.

VFA increased about 42%. The growth was much lower than that in anaerobic biological pre-treatment process (about 300% (Wang et al., 2012)). Although VFA concentration

increased, pH got a little higher after hydrolysis. This may be due to increase of alkalinity (about 27%), which is related to the buffer capacity of the materials. During thermal hydrolysis, a large amount of compounds were released from cells. Many of them had buffering ability and thus contributed to the increase of alkalinity (Phothilangka et al., 2008).

3.3.3 Digestibility

As discussed before, thermal hydrolysis had little impact on inhibitors which directly affect digestion efficiency. However, as dissolved organics, which may be digested into VFA and ammonia quickly, increased after thermal hydrolysis, it's still hard to tell the effect of thermal hydrolysis on digestibility of municipal biowaste.

Consequently, a BMP test monitoring the real digestion ability was carried out. Figure 3.3 gives the specific methane production during the test.

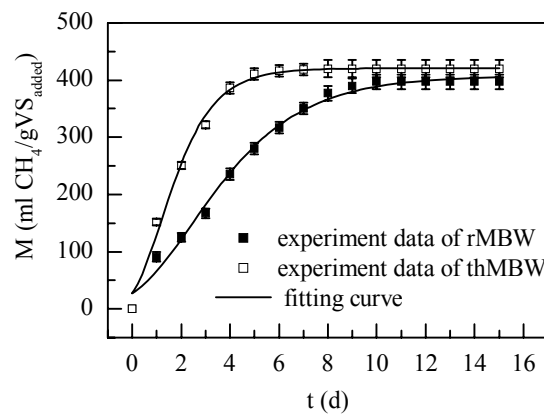


Figure 3.3 Methane production during the BMP test

From these data, maximum methane production rate and BMP can be calculated. Improved Gompertz model (Zwietering et al., 1990; Lay et al., 1998) was used for nonlinear fitting of the results. The model can be expressed as following equation:

$$M = BMP \times \exp \left\{ -\exp \left[\frac{R_{\max} \times e}{BMP} (\lambda - t) + 1 \right] \right\} \quad (3-4)$$

M: accumulated specific methane production, ml CH₄/g VS_{added}; BMP: biochemical methane potential, ml CH₄/g VS_{added}; t: reaction time, d; R_{max}: maximum methane production rate, ml CH₄/(g VS_{added}·d); λ: lag time for methane production, d; e: natural logarithm.

In this experiment, no lag phase was found for both rMBW and thMBW. As a result, λ was taken as 0. Nonlinear fitting of the results was carried out using Origin 8.0. Results were shown in Table 3.4. Some results from other studies were also listed in Table 3.4. Results of this test were acceptable from both R^2 values and comparison with other studies.

Although improvement of BMP was little (about 3%), maximum methane production rate increased 115%. This was higher than the 97% increase in former lab studies (Wang et al., 2009), which might also result from unequal heating by oil bath in the lab studies.

Table 3.4 Results of BMP assay and other studies

	BMP (ml CH ₄ /g VS _{added})	R _{max} (ml CH ₄ /(g VS _{added} ·d)	R ²	Reference
rMBW	407.5	60.7	0.9921	
thMBW	420.7	130.3	0.9924	
food waste	472	-	-	Cho et al., 1995
WSS	151	-	-	Wang et al., 2004

3.4 Effect of thermal hydrolysis on physical properties of MBW

3.4.1 Particle size distribution

Particle size has been proved important to digestion performance of municipal biowaste. Reduction of particle size can increase the specific surface of waste to the microorganisms and thus increase digestion efficiency. Particle size may also affect fluid properties like settling velocity and dewaterability (Palmowski and Müller, 2000).

As a result, particle size distribution of rMBW and thMBW was determined. In this research, sieving method was used. Results were shown in Figure 3.4.

Except for TS, distribution of VS was also listed for distribution of organic solids. After thermal hydrolysis, particles larger than 900 μ m almost vanished. All the other particles larger than 74 μ m also had a certain decrease.

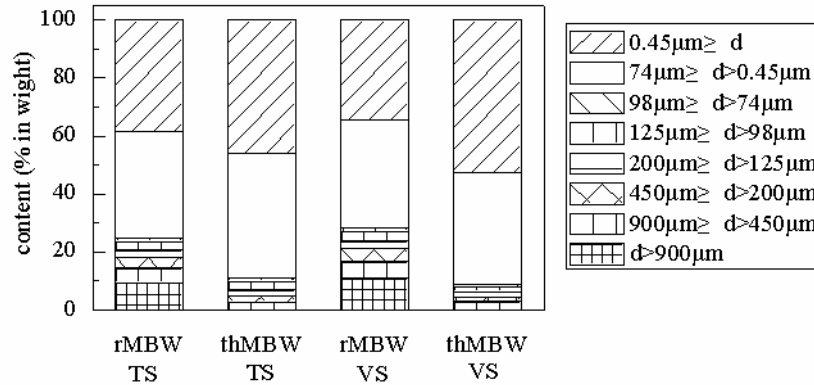


Figure 3.4 Particle size distributions in rMBW and thMBW

Variations of different sizes of particles were listed in Table 3.5. It was found the changes of VS were even larger than TS. This means decrease of organic particulates larger than 74µm were more significant than inorganic ones. Sizes of particles larger than 74µm were diminished into less than 74µm or even into dissolved phase ($\leq 0.45\mu\text{m}$). For TS, substances with size from 0.45µm to 74µm increased 16.5%, while soluble substances ($\leq 0.45\mu\text{m}$) increased 20.5%. However, from VS result, it can be concluded that organic substances from 0.45µm to 74µm increased only 4%, while soluble organics increased 52.9%. This means more organics were dissolved than inorganic matters during thermal hydrolysis.

Table 3.5 TS and VS increase in different sizes after thermal hydrolysis (%)

	TS	VS
d>900µm	-100.0	-100.0
900µm≥d>450µm	-45.2	-50.0
450µm≥d>200µm	-46.5	-62.0
200µm≥d>125µm	-10.3	-40.3
125µm≥d>98µm	-2.2	-35.6
98µm≥d>74µm	-13.1	-57.5
74µm≥d>0.45µm	16.5	4.0
0.45µm≥d	20.5	52.9

As more organics were dissolved or size-reduced, their contact with microorganisms could increase. Consequently, it could contribute to improvement of digestibility. However, it

may have some adverse effect on fluid properties of the waste. As a result, fluid properties like viscosity, dewaterability, and settling performance of the waste were determined. The results will be discussed below.

3.4.2 Viscosity

From the results of particle size distribution, it was found particles in smaller sizes increased after thermal hydrolysis. This higher number of smaller particles may result in more particle-particle interactions and higher viscosity. However, as shown in Table 3.6, viscosity of municipal biowaste decreased sharply after thermal hydrolysis. This decrease in viscosity means better mobility of the waste. Energy cost for pumping in the following disposal processes could be diminished consequently.

Table 3.6 Fluid properties of rMBW and thMBW

Indicators		rMBW	thMBW
Viscosity (mPa·s)		7000-80000	150-2150
Dewaterability	Dryness of dewatered cake (%)	16-38	46-62
	CST (s)	1213-1335	347-353
Settling velocity (cm/h)		0 (not observed)	6.5-9.3

Decrease of viscosity can partly be explained by Krieger-Dougherty Model (Krieger and Dougherty, 1959).

$$\eta = \eta_0 \left(1 - \varphi / \varphi_m\right)^{-[\eta]\varphi_m} \quad (3-5)$$

where η is the viscosity of suspension, η_0 is the viscosity of base medium, which is water in this case, φ is the volume fraction of solids in the suspension, φ_m is the maximum volume fraction of solids in the suspension and $[\eta]$ is the intrinsic viscosity of base medium, which is 2.5 for spheres (Krieger and Dougherty, 1959).

According to equation (3-5), viscosity could be reduced from both decrease of φ and increase of φ_m .

The parameter ϕ_m , could be increased due to change of particle size distribution. As discussed before, particle sizes decreased after thermal hydrolysis. It means the bulk density of these particles could be increased and also the value of ϕ_m .

The parameter ϕ , could be decreased due to the dilution effect of steam. As the MBW was diluted in pilot-scale thermal hydrolysis, ϕ got lower and caused a lower viscosity. However, it may not be the main reason for viscosity decrease as similar results were also found in lab studies. Table 3.7 gives results of lab studies, in which MBW were heated by oil bath and thus no dilution happened. It was found the viscosity also decreased a lot.

Besides of the above two reasons, solubilisation of solids and destruction of large colloid may also contributed to the decrease of viscosity. However, their effects were not described in Krieger-Dougherty' model. However, these changes may be more important. For example, extracellular polymeric substances (EPS), which contributes a lot to the viscosity of WSS (Magara et al., 1976), can be broken into small molecules during thermal hydrolysis.

Table 3.7 Viscosity of municipal biowaste before/after thermal hydrolysis (Liu et al., 2012)

	before	after
RKW	36000 ± 1414	1658 ± 177
FVW	6250 ± 1768	663 ± 53
WSS	13,500 ± 500	1625 ± 331

3.4.3 Dewaterability

As listed in Table 3.6, thermal hydrolysis also changed dewatering performance of municipal biowaste.

Two parameters were used for evaluation of dewaterability. One is dryness of dewatered cake, which was got from filter pressing under 1 MPa of pressure for 40 min. Another one is CST, which can be easily tested and thus has been widely used for evaluation of dewaterability. It was found dryness of dewatered cake increased from 16-38% to 46-62%, while CST decreased from about 1300s to 350s. Both results indicated better dewaterability for thMBW.

Dewaterability is an important parameter for treatment of biowaste. Improved dewaterability indicated reduction of dewatered cake, resulting in less cost for transportation and post-disposal.

3.4.4 Settling performance

As shown in Table 3.6, no obvious solid settlement was found for rMBW. While for thMBW, settling velocity of 6.5-9.3 cm/h was found for the solid-liquid separating surface.

Settling performance of solids is affected by both particle size and viscosity. Stokes's Law (Equation 3-6) described their relationship for gravitational settling of spherical particles.

$$v_s = \frac{2(\rho_p - \rho_f)}{9\mu} gR^2 \quad (3-6)$$

where v_s : particle's settling velocity, m/s; ρ_p : mass density of the particle, kg/m^3 ; ρ_f : mass density of the fluid, kg/m^3 ; g : gravitational acceleration, m/s^2 ; R : radius of the particle, m; μ : dynamic viscosity, N s/m^2 .

From this equation, it's clear that gravitational settling velocity decreases with increasing viscosity and decreasing particle size. Despite of particle size decrease, settling performance was improved by thermal pre-treatment due to decrease of viscosity. This improvement indicated that solid-liquid separation could be used in the following anaerobic digestion processes to improve digestion efficiency.

3.5 Conclusion

In this research, thermal hydrolysis was applied to municipal biowaste at 175°C for 1 hour. Effect of thermal hydrolysis on biochemical and physical properties of municipal biowaste was discussed in this chapter. It was found:

1) Thermal pre-treatment resulted in VSS hydrolysis ratio of 38.3%, and thus increased VDS/VS ratio from 0.19 to 0.48, and sCOD by 56%. These results were a little higher than those obtained in lab studies, which may result from unequal heating in the lab studies.

2) Thermal pre-treatment increased concentrations of VFA by 42%, while no significant variation was found for ammonia and pH. This may result from increasing buffering ability of the biowaste.

3)BMP test found digestibility of municipal biowaste was improved by thermal hydrolysis, with a little improvement on ultimate methane production and 115% increase on maximum methane production rate, comparing with 97% in lab studies.

4)Thermal hydrolysis decreased particles larger than 74 μ m, while substances from 0.45 μ m to 74 μ m and soluble organics (\leq 0.45 μ m) increased. More organics were found dissolved after thermal hydrolysis than inorganic matters, corresponding with the results of hydrolysis ratio. Decrease of particle size could contribute to improvement of digestibility. However, it may have some adverse effects on fluidity of the waste.

5)Viscosity decreased after thermal pre-treatment, while dewatering and settling performance were improved. This was resulted from interaction of molecule destruction, particle size variation, dilution, and so on. Especially, improvement on settling performance indicated that application of solid-liquid separation become possible.

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Chapter 4 Effect of thermal hydrolysis on anaerobic digestion of municipal biowaste

4.1 Introduction

Many researchers have proved that thermal hydrolysis can improve digestibility of the sewage sludge (Fisher and Swanwick, 1971; Haug et al., 1978). Former lab studies (Wang et al., 2009; Liu et al., 2012) and the results in Chapter 3 also proved its improvement on digestibility of mixed MBW. Most of these present conclusions were based on batch experiments like BMP tests, while little studies were carried out in continuously operated reactors. However, as batch and continuous experiments have different operating conditions, both substrates consumption and microorganism growth could be different. These differences may affect the effect of thermal hydrolysis on digester performance.

According to this, effect of thermal hydrolysis was evaluated in continuously operated digesters in this Chapter. Two pilot-scale digesters, fed with rMBW and thMBW respectively, were compared for their stability and digestion performance. Moreover, batch monitoring was also carried out during operation of these continuous reactors, which can provide better description about the effect of thermal hydrolysis on the digestion processes.

4.2 Materials and methods

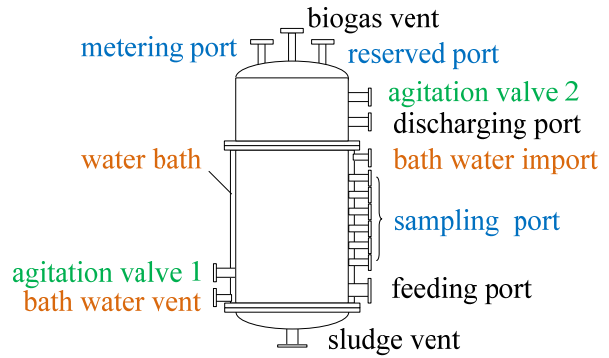
4.2.1 Digesters

Pilot-scale digesters were used in this research, fed with rMBW and thMBW separately. The digesters were 1 m tall and 0.5 m in diameter, with an effective volume of 175l. The digesters were kept in mesophilic state by water bath, with temperature ranged from 35°C to 37°C. Hydraulic agitation was used for mixing, with strength of 5 min per hour. The agitation system was combined by two agitation valves, a screw pump, and an automatic control system. Digestate was pumped from the bottom of the digester (through agitation valve 1), and returned into the digester from agitation valve 2 in the top of digester. The pump had a rated flow of 5 m³/h, which was sufficient to pump all the digestate in 5 min.

Figure 4.1 gives photo image and sketch of the digesters.



(a)



(b)

Figure 4.1 Pilot-scale anaerobic digesters: (a) photo image; (b) sketch

4.2.2 Operating parameters

Digesters were operated in a semi-continuous mode, with daily feeding and discharging. Two digesters were used for this experiment, one fed with rMBW and the other with thMBW. Two operations were run for each digester, with organic loading rates (OLRs) of about 1.5 and 3 VS/(m³·d). They were comparable with or higher than that in anaerobic digestion of sewage sludge, which usually is 1-2 kg VS/(m³·d). Each operation was run for about 2-3 times of HRT to reach a stable state.

Detailed operating parameters were listed in Table 4.1. The actual OLRs had some difference between the two digesters, as feed properties varied from average values during the experiments. HRTs were also different, due to different solid concentration between rMBW and thMBW.

Table 4.1 Operating parameters of digesters treating rMBW and thMBW

operation	r-1	t-1	r-2	t-2
feed type	rMBW	thMBW	rMBW	thMBW
feeding flow (kg/d)	3	5	7	10
VS conc. in feed (g/kg)	83.1 ± 13.8	61.7 ± 9.1	80.7 ± 9.4	50.2 ± 11.2
OLR (kg VS/(m ³ ·d))	1.4 ± 0.2	1.8 ± 0.3	3.2 ± 0.4	2.8 ± 0.6
HRT (d)	58	35	25	18
operating time (d)	106	120	80	66

4.2.3 Sampling and analysis

Volume of biogas produced from the reactors was detected by flow meters. A record was made for daily biogas production.

Sampling and analysis of pH, 4S, sCOD, and VFA were done weekly, while analysis of methane contents, ammonia, and alkalinity were done for twice weeks a time. Analysis methods were as same as written in section 3.1.3.

4.2.4 Batch monitoring during one treatment cycle

Except for weekly monitoring of effluent, determination was also carried out during one treatment cycle when digesters reached a stable status (r-1 at day 95, t-1 at day 118, r-2 and t-2 at day 186). Change of these parameters revealed digester stability and performance as well as weekly data of effluent properties.

Batch monitoring of pH, sCOD, VFA, biogas production, and methane content was carried out inside one treatment cycle from feeding to discharging (24 hours in this research). Samples were collected soon after 5 minutes' agitation each time. Sampling was set at the time of 0 (right before feeding), 0.33, 0.67, 1, 1.5, 2, 2.5, 3, 3.5, 4, 5, 6, 7, 8, 10, 15, 19, and 24 hours. Less sampling was carried out with time going, as reaction was not so intense at that time.

4.3 Digester stability

VFA and pH are important parameters indicating digester stability. Optimal pH range of 6.8-7.2 and VFA lower than 2000 mg/l was usually suggested for stable operation of anaerobic digesters (McCarty and McKinney, 1961).

As the digesters were operated in a semi-continuous mode, effluent properties cannot represent the status of digesters all the time. While pH and VFA concentrations of effluent are inside optimal ranges, their values might exceed these ranges before discharging. As a result, not only did the effluent properties, but also monitoring during one treatment cycle (24 hours) were discussed.

4.3.1 Effluent properties

Effluent of anaerobic digesters, known as digestate in some cases, should be treated further before final disposal. Digester effluent in a stable status ensures stable utilization or operation in the following disposal processes.

1)pH

One of the major indicators of digester stability is pH. During anaerobic digestion, the processes of hydrolysis and acidification produce acids and thus pH decrease. While in the process of methanogenesis, acids are utilized to produce biogas, which resulted in an increasing pH. The value of pH is affected by different buffers inside the digestate. An optimal pH range of 6.8-7.2 was suggested by McCarty and McKinney (1961), while a wider range of 6.5-8 was also acceptable (Capri and Marais, 1975; Brummeler and Koster, 1989).

Figure 4.2 gives variation of pH in the effluent. As the two digesters started operation from different conditions, the initial pH value was quite different. In operation t-1, an initial pH of 6.87 was found. It was relative lower but still in the optimal range. This low pH was originated from a short HRT of only 9.5 d in the former operation, while HRTs longer 10 d were usually suggested for anaerobic digestion. Despite of the low initial value, effluent pH of t-1 increased with time going. After about 40 d' operation, pH value of effluent in both r-1 and t-1 reached a relative stable status, with an average value of 7.62 and 7.48, respectively. With increased OLR, effluent pH increased. In r-2 and t-2, average effluent pH of 7.69 and 7.55 was found.

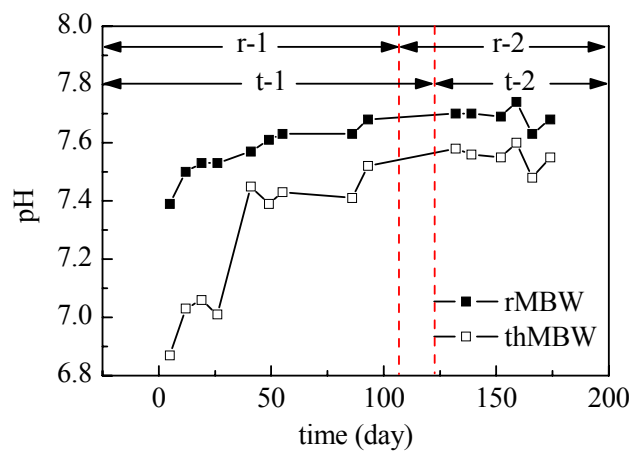


Figure 4.2 Effluent pH of digesters treating rMBW and thMBW

Under both of the two OLRs, variation of effluent pH was not so significant after reaching a stable status. However, with OLR increasing, pH values of the digesters also increased. This can be explained by the results of alkalinity listed in Table 4.2. Alkalinity means the buffer ability of a solution/mixture to the acids. Although more acids may be produced at a higher OLR, their effect to pH variation was buffered by increasing alkalinity.

Table 4.2 Alkalinity of digesters treating rMBW and thMBW

	r-1	t-1	r-2	t-2
Alkalinity (mg CaCO ₃ /l)	11129± 741	6349±852	18644 ± 1185	8065±954

Results of alkalinity can be also used to explain the higher effluent pH from the digester treating rMBW than that treating thMBW. As the feed solid concentration of feeding materials was higher in rMBW than that in thMBW, alkalinity values in digester treating thMBW were relatively lower. Alkalinity values in the former digester were almost double of those in the latter. This means, the digester fed with rMBW had better buffering ability against VFA production.

2)VFA

VFA are important intermediate products in the process of anaerobic digestion. They are produced from the process of acidification, and are consumed in the step of methanogenesis to produce biogas. Although used as feed stock, VFA at high concentrations can also inhibit the activity of methanogenus. For stable operation of anaerobic digesters, VFA should be kept in a proper range. Commonly, inhibition is supposed to happen when VFA concentrations are higher than 2000 mg/l (Kroeker et al., 1979; McCarty and McKinney, 1961).

VFA concentrations in the effluent of both digesters were shown in Figure 4.3. For the digester treating thMBW, VFA concentrations were found relatively lower than that of rMBW in both operations. For the digester treating rMBW, VFA accumulation was found with OLR increase. VFA concentration increased from about 170 mg COD/l to 700 mg COD/l in this digester, while no significant accumulation was found for the digester treating thMBW.

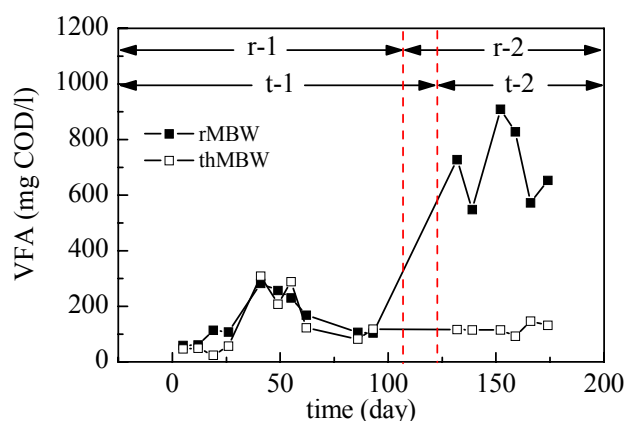


Figure 4.3 VFA in the effluent of digesters treating rMBW and thMBW

3)VFA/alkalinity

Some researchers have found that VFA inhibition only comes from un-ionized VFA, while ionized VFA has little inhibition (Mata-Alvarez and Llabrés, 1988; Mata-Alvarez, 2002). Ionization of VFA depends on pH. More VFA are ionized at a higher pH. While at the same time, pH is affected by the buffer ability of the mixture. As a result, some researchers brought out a combined indicator, VFA/alkalinity ratio, for evaluation of system stability during anaerobic digestion. Mosquera-Corral et al.(2001) found VFA/alkalinity ratio higher than 0.3 lead to unstable operation of anaerobic digestion. Similar result was also found by Borja et al. (2004), suggesting an upper limitation of VFA/alkalinity ratio from 0.3 to 0.4.

VFA/alkalinity ratio was calculated from average VFA and alkalinity concentrations. The results were listed in Table 4.3.

Table 4.3 VFA/alkalinity ratio of digesters treating rMBW and thMBW

	r-1	t-1	r-2	t-2
VFA (mg COD /l)	171±59	120±19	706 ± 143	159±56
Alkalinity (mg CaCO ₃ /l)	11129± 741	6349±852	18644 ± 1185	8065±954
VFA/alkalinity ratio	0.015	0.019	0.038	0.020

As shown, VFA/alkalinity ratios were kept under 0.04 at all the four operations, indicating stable operation under these situations. VFA/alkalinity ratio increased with OLR due to increased VFA. But, variation was more significant in rMBW treatment.

As a conclusion, it can be said, although the digester treating thMBW had lower pH and alkalinity values, it had a better buffer ability to both VFA concentrations and VFA/alkalinity ratio with increasing OLR.

4.3.2 Batch monitoring during one treatment cycle

As discussed before, monitoring of pH and VFA was also carried out from feeding to discharging for each operation.

As shown in Figure 2.8, in anaerobic digestion processes, organic acids are formed after hydrolysis and acidogenesis, resulting in decreasing pH. These acids are converted into acetic acid in the stage of acetogenesis. Value of pH is then recovered in the process of methanogenesis, in which acetic acids are utilized to produce biogas (Rittmann and McCarty, 2001). Base on this, anaerobic digestion of municipal biowaste is simplified as Figure 4.4 for following discussion.

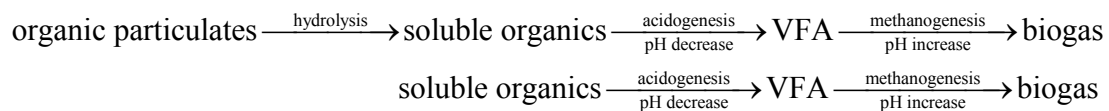


Figure 4.4 Anaerobic digestion process (simplified)

Variations of pH and VFA during one treatment cycle were shown in Figure 4.5. Acetic acid and propionic acid were dominant acids in the VFA and their concentrations were also graphed in Figure 4.5. Similar pattern was found for variation of total VFA, acetic acid, and propionic acid.

It can be found in all the four operations, pH and VFA were still in a suitable range. Highest VFA concentration and lowest pH was found in operation t-2, with the values of 868 mg/l and 7.48, respectively. Detailed analysis of these data was as following.

For the digester treating thMBW, VFA rose quickly after feeding with decreasing pH. Highest VFA concentration reached to about 800 mg/l and 900 mg/l in operation t-1 and t-2, accompanied with pH decrease to 7.57 and 7.48. With a higher OLR, accumulation of VFA got quicker. For t-1, it took about 2 hours to reach the peak point, while only 1 hour cost for t-2. After that, VFA concentrations decreased gradually with increasing pH. After about 8 hours, values of pH and VFA concentrations recovered to the initial levels.

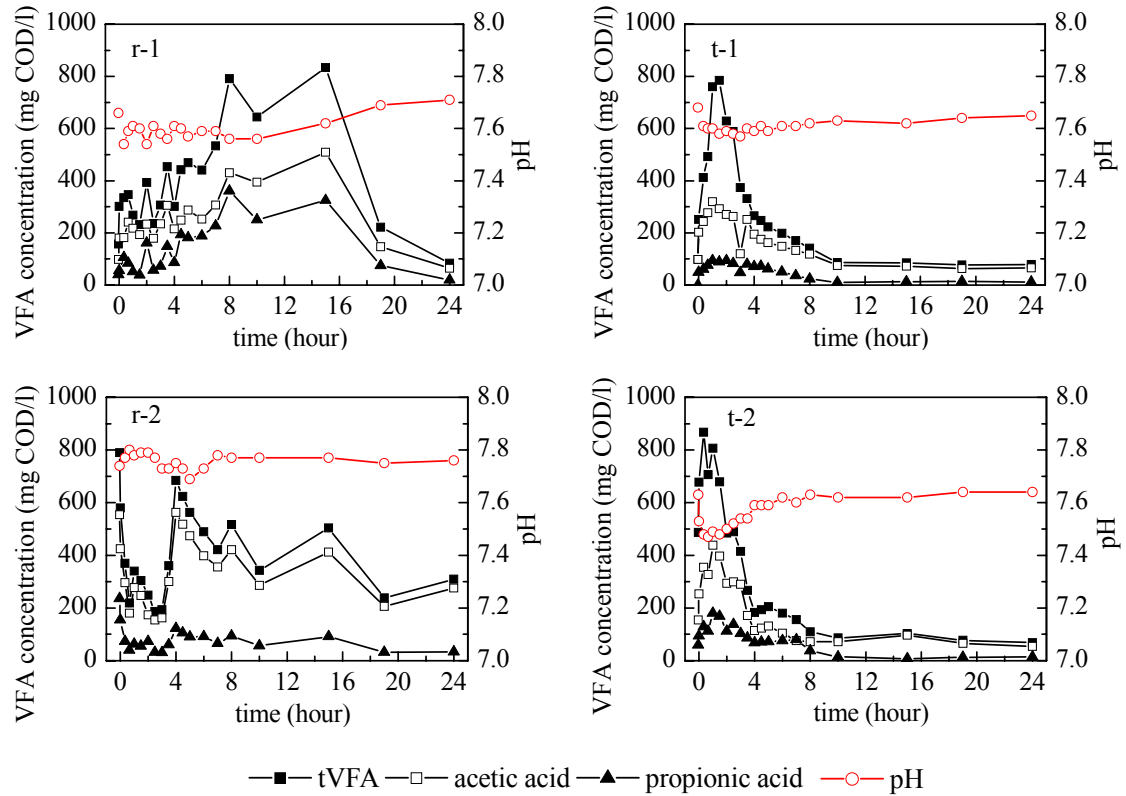


Figure 4.5 Changes of VFA and pH during one treatment cycle (24h)

For the digester treating rMBW, there was a lag phase before VFA rising, lasting about 6 hours for r-1 and 4 hours for r-2. After that, VFA increased to peak values of about 800 mg/l and 700 mg/l in operation r-1 and r-2, accompanied with pH decrease to 7.56 and 7.69. Then they were kept in a relative high level for 10 hours in r-1 and 12 hours in r-2. VFA concentrations recovered until the last 4 hours for both operations. Contrary variation of pH can be found in both operations.

Comparing the results of the two digesters, it's clear that change of pH and VFA was less and slower in rMBW digestion than that in thMBW digestion. This probably resulted from both insufficient mixing and hydrolysis limitation. Without thermal pre-treatment, solid hydrolysis became limiting step in rMBW digestion. As shown in Figure 4.4, formation of soluble organics from solid substrates was limited and thus no accumulation of VFA occurred at the first several hours. With time going, more organic particulates were hydrolyzed and thus hydrolysis was no longer the limiting step. When thermal pre-treatment was used,

hydrolysis was no longer the limiting step. As a result, VFA in operation t-1 and t-2 increased quickly.

Combined with the results of effluent properties, it can be concluded that with thermal pre-treatment, lower pH and alkalinity will be found for the digester. When thermal hydrolysis was applied, solid hydrolysis was no longer the limiting step and more soluble organics were contained in the feeding materials. As a result, when digesters are operated by a semi-continuous mode, a large amount of VFAs may be produced soon after feeding under a shock organic loading. Digesters with thermal pre-treatment would get unstable more easily than that without thermal pre-treatment. However, this situation can be relieved if continuous feeding and discharging mode are applied instead of the semi-continuous operation.

Moreover, without thermal pre-treatment, digestion efficiency was limited by solid hydrolysis and thus accumulation of VFA could also happen. As the situation in r-2, as solid hydrolysis limited the process of digestion in the first several hours, VFA was not sufficiently degraded and accumulation was found. VFA concentrations in the effluent were much higher than those of the other operations. This implied that if shorter HRT are chosen, the digester treating rMBW may get unstable rather than that with thermal pre-treatment.

4.4 Digester performance

Digester performance will be discussed from the results of biogas production, methane content, and removal of substrates.

4.4.1 Biogas and methane production

Biogas production of the two digesters were recorded daily and plotted in Figure 4.6. Average values of daily biogas production, specific biogas production per volume and biogas yield were calculated and listed in Table 4.4. It can be found biogas yield declined with increasing OLR. However, obvious difference was not found for the two digesters.

Average methane content, specific methane production and methane yield in each operation were also calculated in Table 4.4. It can be found, content of methane in biogas was higher in thMBW digestion. Due to this, specific methane productions and methane yields in

the digester treating thMBW were higher than that treating rMBW in both operations. However, these differences were not obvious as well.

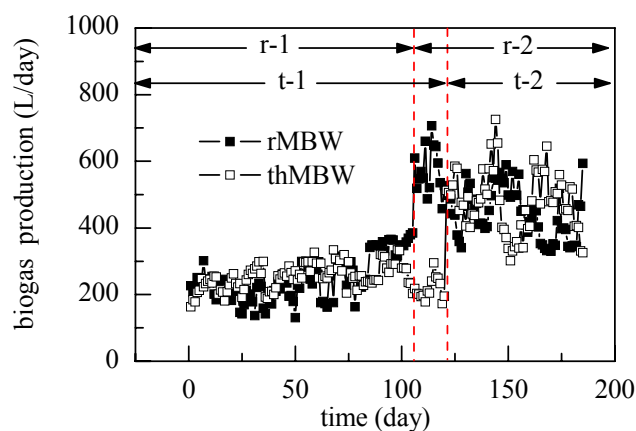


Figure 4.6 Daily biogas productions of digesters treating rMBW and thMBW

Table 4.4 Biogas production of digesters treating rMBW and thMBW

	r-1	t-1	r-2	t-2
Daily biogas production (l/d)	250 ± 64	260 ± 38	469 ± 90	476 ± 91
Specific biogas production (l/(l·d))	1.4 ± 0.4	1.5 ± 0.2	2.6 ± 0.5	2.6 ± 0.5
Biogas yield (l/kg VS)	854 ± 220	843 ± 123	775 ± 149	771 ± 148
Methane content (%)	68.0 ± 1.7	69.2 ± 0.3	65.9 ± 3.0	69.6 ± 0.0
Methane yield (l/kg VS)	544 ± 140	566 ± 82	499 ± 96	517 ± 99

4.4.2 Removal of VS

VS was used to represent the amount of organic substrates. Its average concentration in both feed and effluent were listed in Table 4.5. Results were similar to those from biogas and methane production. Significant difference was not found for the two digesters, while VS removal rate was a little higher in thMBW digestion.

Table 4.5 VS removal of digesters treating rMBW and thMBW

	r-1	t-1	r-2	t-2
VS in feed (g/kg)	83.1 ± 13.8	61.7 ± 9.1	80.7 ± 9.4	50.2 ± 11.2
VS in effluent (g/kg)	24.0 ± 1.8	19.3 ± 1.8	35.5 ± 1.0	21.6 ± 4.5
VS removal rate (%)	71	69	56	57

4.4.3 Methane production

Although there was little difference between the two digesters in digestion performance, significant difference was found during monitoring of one treatment cycle. Biogas and methane productions inside one treatment cycle were plotted in Figure 4.7. It can be found, the biogas production rates were faster in thMBW digestion.

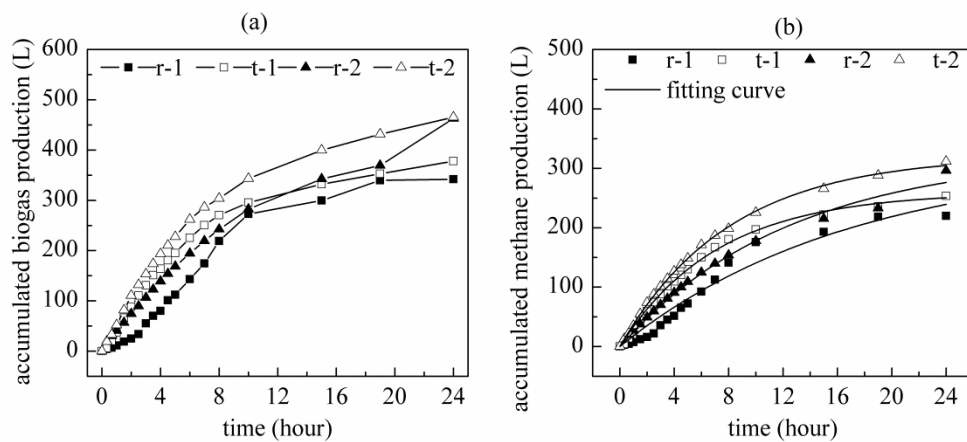


Figure 4.7 Accumulated biogas production (a) and methane production (b) inside one treatment cycle

Variations of methane content during a treatment cycle were listed in Figure 4.8. Average methane content was calculated combining the data of biogas production. The average methane content for r-1, t-1, r-2, and t-2 in this treatment cycle was 67.9%, 68.9%, 64.0%, and 69.6%. These results were close to those from the weekly monitoring results and were believed acceptable.

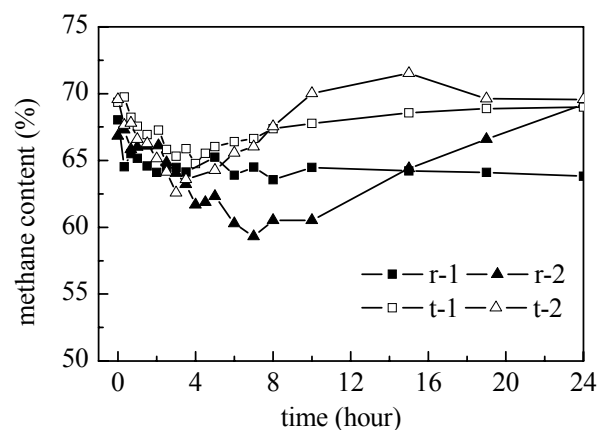


Figure 4.8 Methane content inside one treatment cycle

Biogas produced during anaerobic digestion is mainly composed by methane and carbon dioxide. Methane is generated from acetoclastic reaction which utilizes acetic acid as carbon

source (as shown in Equation 4-5), and hydrogenotrophic methanogenesis which uses carbon dioxide as carbon source (as shown in Equation 4-6).



Carbon dioxide is produced in reaction (4-5) and consumed in reaction (4-6). Moreover, part of the carbon dioxide is also produced from the stage of acidogenesis (Rittmann and McCarty, 2001).

From Figure 4.8, reduction of methane content was found for all operations after feeding. This can be explained by production of carbon dioxide from acidogenesis. More carbon dioxide was produced than methane, resulting in declined methane content. When reaction went on, carbon dioxide was utilized by hydrogenotrophic methanogens to produce methane according to Equation 4-6. As a result, carbon dioxide content declined and methane content increased.

In the process of thMBW digestion, recovery of methane content was faster than that in rMBW digestion, which corresponded with the variation of pH and VFA. It resulted from improved reaction rates of hydrolysis and following acidogenesis step by thermal hydrolysis. After thermal hydrolysis, production of carbon dioxide from acidogenesis got quicker, while reactions producing methane also speeded up.

By multiplying methane production and average methane content between two sampling time, accumulated methane productions were calculated and plotted in Figure 4.7 (b).

Although thermal pre-treatment improved solid hydrolysis of the biowaste, it can be calculated that more than 50% of the organics still existed in the form of solid and less than 10% of total COD were from VFAs. Anaerobic digestion of thMBW may still be rate-limited by hydrolysis. As no significant inhibition was all the four operations, anaerobic digestion was assumed to be rate-limited by hydrolysis and the reaction was assumed as first-order based on the model established by Eastman and Ferguson (1981). Substrate concentration can be calculated as:

$$S_h = S_{h,0} \cdot \exp(-k_h \cdot t) \quad (4-7)$$

where S_h means concentration of substrates that can be hydrolyzed at the time of t ; $S_{h,0}$ means initial concentration of substrates that can be hydrolyzed; t means reaction time; k_h means hydrolysis rate.

Production of methane can then be expressed as:

$$P_m = k \cdot (S_{h,0} - S_h) = k \cdot S_{h,0} \cdot (1 - \exp(-k_h \cdot t)) = a \cdot (1 - \exp(-k_h \cdot t)) \quad (4-8)$$

where P_m means accumulated methane production at the time of t ; k is constant related to specific methane production of the substrate ; a is constant multiplied by k and $S_{h,0}$.

Simulation was carried out using data in Figure 4.7 (b). Results were listed in

Table 4.6. The simulation results showed the former assumption of hydrolysis limitation was reasonable with high R-square values. Although the final biogas productions were similar for rMBW and thMBW digestion, their reaction rates were quite different. Hydrolysis rates in thMBW digestion were about double of that in rMBW digestion. It also corresponded with the former BMP test results, in which a 115% increase was found for specific biogas production rate when thermal hydrolysis was used.

Table 4.6 Simulation of methane production following first-order hydrolysis model

	k_h (d ⁻¹)	Adj. R-Square
rMBW (r-1)	1.42 ± 0.31	0.9577
thMBW(t-1)	3.31 ± 0.12	0.9965
rMBW (r-2)	1.90 ± 0.14	0.9913
thMBW(t-2)	2.98 ± 0.05	0.9991

It can be concluded from these results, although thermal pre-treatment cannot hydrolyze all the solid particulates into soluble phases, the residual particulates were easier to be digested. One of the major reasons could be broken-up of big particulates into small ones and release of biodegradable particulates from cells during thermal hydrolysis (Phothilangka et al., 2008). These smaller or released particulates are much easier to be degraded than their original forms. With faster reaction rate, digesters fed with thMBW could be operated under a shorter HRT and thus digester volume can be saved.

4.5 Discussion on energy consumption

From the above results, it was found although HRTs were shorter, more VS were removed and more bio-energy was recovered when thermal hydrolysis was applied as pre-treatment before anaerobic digestion. However, as thermal hydrolysis is usually operated at a high temperature, its application may be limited by the high energy input used for heating. Optimal temperature was chosen in this research as 175°C, which is the same with the results from Haug (1978) and similar to the temperature range of 150-170°C that used in Cambi™ and THELYS™ (Cambi™, 2012; Veolia Water Solutions & Technologies, 2012). Dwyer (2008) found decrease of temperature in some ranges had little effect on anaerobic digestibility of activated sludge, but only to a value of 140°C. As a result, energy consumption is still worth consideration when using thermal hydrolysis as pre-treatment before anaerobic digestion.

Parts of this energy consumption can be balanced by improved bio-energy recovery. Moreover, improvement on effluent fluidity by thermal hydrolysis can also contribute to energy balance of the whole system. Table 4.7 listed results of viscosity and dewaterability of effluent in the four operations.

Table 4.7 Viscosity and dewaterability of effluent from digesters treating rMBW and thMBW

	r-1	t-1	r-2	t-2
Viscosity (mPa·s)	1114 ± 403	6 ± 1	7583 ± 1393	11 ± 3
Dryness of dewatered cake (%)	17 ± 6	37 ± 1	not detected	not detected
CST (s)	not detected	not detected	2896 ± 60	191 ± 5

Effluent viscosity in thMBW digestion was much lower than that in rMBW digestion. As a result, heat and mass transfer of the digestate can be improved, which help with energy saving as well (El-Mashad et al., 2003). Energy can also be saved in the process of pumping (Moeller and Torres, 1997). Dewatering performance of effluent was also better with thermal pre-treatment. As effluent of digesters treating municipal biowaste had high concentrations of solids, further disposal were needed. With improved dewaterability, energy cost for dewatering, transportation, and advanced disposal can be also reduced.

4.6 Conclusion

In this Chapter, effect of thermal hydrolysis was evaluated in pilot-scale continuous digesters. Digesters treating rMBW and thMBW were compared for their stability and performance under OLRs of about 1.5 and 3 kg VS/(m³·d). Results from both continuous operation and batch monitoring were used for evaluation. Following conclusions were obtained:

1) Thermal hydrolysis improved digester stability, especially under a higher OLR of about 3 kg VS/(m³·d). With thermal hydrolysis, VFA concentrations in the digestate kept at about 120 mg COD/l and VFA/alkalinity ratio was about 0.02. While for the digester without thermal hydrolysis pre-treatment, when OLR increased, VFA concentrations in the digestate increased to about 700 mg COD/l and VFA/alkalinity ratio also increased to 0.04.

2) The batch monitoring revealed that without thermal pre-treatment, formation of VFA was lagged by limitation from solid hydrolysis. As a result, VFA was not sufficiently degraded and was accumulated. This implied that if shorter HRT are chosen, the digester without thermal pre-treatment may get unstable easier.

3) From the continuous operation results, obvious difference was not found for digestion performance of both rMBW and thMBW, although the methane production and VS removal was a little higher in thMBW digestion. Despite of this, batch monitoring results during a treatment cycle showed thermal pre-treatment promoted hydrolysis of solids and thus increased reaction rate. This implied better performance would be achieved with thermal hydrolysis pre-treatment when reaction time is shortened under a lower HRT.

4) Methane contents in thMBW digestion were higher than that in rMBW digestion. The variations of methane contents during batch monitoring were also greater in thMBW digestion, which resulted from improved reaction rates of hydrolysis and following acidogenesis step by thermal hydrolysis.

5) Thermal hydrolysis pre-treatment costs energy for heating. However, this cost can be recovered to some extent from increased bio-energy recovery, decreased viscosity, and improved dewaterability of the digestate.

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Chapter 5 Application of anaerobic sequencing batch reactors for the digestion of thermally hydrolyzed municipal biowaste

5.1 Introduction

Thermal hydrolysis pre-treatment was found not only directly hydrolyzing solids into soluble phases, but also increasing reaction rate in the following digestion process from results in Chapter 3 and 4. However, it was calculated that, more than 50% of VS remained in the form of particulates and VFAs occupied less than 10% of the total COD in thMBW. Anaerobic digestion of thMBW might still be rate-limited by hydrolysis. As a result, measures should still be taken to improve hydrolysis in order to achieve higher digestion efficiency.

According to this, ASBR was used for digestion of thMBW in this research. ASBR has a settling phase inside one treatment cycle. Solid substrates and microorganisms can be settled in this phase, thus SRT can be separated from HRT. Consequently, ASBR can achieve better effluent quality and higher digestion efficiency. Moreover, ASBR can easily be switched from CSTR digester without changing the reactor structure. As a result, cost can be saved for building and operating, especially when applying to a present anaerobic digestion plant.

Many studies of ASBR have been carried out. However, most of them were limited to wastewater treatment. Application of ASBR on MBW digestion was limited. The major block is the high solid concentration of MBW. Solid settlement velocity decreases with increasing solid concentration (Giokas et al., 2003) and thus affects the performance of ASBR. As a result, only a few studies used ASBR to treat high-solid-content MBW (Hur et al., 1999; Luo et al., 2009; Wang W. et al., 2009) and no pilot- or full-scale studies were found by now.

In this study, municipal biowaste with TS of about 110 g/L were treated. It's not able to realize solid-liquid separation under this condition. However, obvious solid-liquid separation was found after thermal hydrolysis pre-treatment. As a result, ASBR was able to be used for anaerobic digestion of thMBW. Lab-scale studies have been carried out by fellow researchers in Prof. Wang's group (Wang W. et al., 2009; Wang Z.J. et al., 2009) and this is the first try in pilot-scale.

This chapter was to check the possibility of using ASBR for MBW digestion. Solid settling behaviours of high-solid-content MBW digestate was studied to find their rules. ASBR performance was compared with that of CSTR to prove its advantages.

5.2 Materials and methods

5.2.1 Digesters

Pilot-scale ASBR and CSTR were used in this study. The structure of ASBR was as same as the CSTR shown in Figure 4.1. Difference between these two types of digesters was the operating method. In ASBR, one treatment cycle (24 hours) was combined by 0.5 hours for discharging and feeding, 11 hours for reaction with 5 minutes/hour hydraulic mixing, and 12.5 hours for reaction and settling (without mixing). The time setting of each phase in one cycle was based from operating experiences in the pilot plants.

5.2.2 Operating parameters

ASBRs and CSTRs were operated under 4 different conditions. The operating parameters were shown in Table 5.1. Results of operations C-1 and C-4 have been used for discussion in Chapter 4. To compare the results of ASBR and CSTR easily, their results is also listed in this chapter.

Table 5.1 Operating parameters of ASBR and CSTR

	1	2	3	4
ASBR	A-1	A-2	A-3	A-4
HRT (d)	35	18	9.5	18
OLR (kg VS/(m ³ ·d))	1.8±0.3	1.7±0.2	1.6±0.1	3.4±0.4
operating time (d)	360	54	84	120
CSTR	C-1	C-2	C-3	C-4
HRT (d)	35	18	9.5	18
OLR (kg VS/(m ³ ·d))	1.8 ± 0.3	1.7 ± 0.1	1.2 ± 0.2	2.8±0.6
operating time (d)	120	54	84	66

For operation 2 and 3, feed materials were diluted by tap water to meet the requirement of different HRTs. Digesters were fed with 5 kg of thMBW every day in operation 1-3 and 10 kg in operation 4. The thMBW was diluted with 5 kg tap water in operation 2 and 15 kg tap water in operation 3. Due to the variation of feed properties, the OLRs in each operation had some difference.

5.2.3 Sampling and analysis

Sampling was carried out twice a week for A-3 and C-3 and weekly for other operations. For operation A-3, sampling was once broken up for about 1 month due to the Spring Festival in China.

Analysis methods were the same as listed in Section 3.2.3 and Section 4.2.3.

5.3 Solid settling in ASBR

5.3.1 Settling theory

Little studies were carried out on the settling characters of ASBR digestate (Zheng, 2003). But many similar studies have been done on settlement of activated sludge. Classical settling process theory has been developed, with four types of settling processes divided:

1)Free settling: This happens in non-flocculent dilutes. Every particle settles independently without mutual interruption. This process can be expressed by Newton's second law and Stoker's law. Settlement of grit in grit chamber and low-solids-content sewage in primary sedimentation tank can be described in this type.

2)Flocculating settling: In flocculating settling, suspended solid concentrations are not so high (usually with SS concentration of 50-100 mg/L). But particles can flocculate and their sizes thus increase as they settle. During the settling process, weight, shape and settling velocity of the particles keep changing. As a result, it's hard to express this process by theoretic equations. Settling of activated sludge in secondary sedimentation tank follows this pattern and the settling behaviors are judged by experiments.

3)Zone Settling /hindered Settling: At SS concentration higher than 5000 mg/L, particles' settling is interrupted by other particles. Their relative location is not changed and settles as a whole. Clear solid-liquid interface can be observed during settling. Solid settlement at the

bottom of secondary sedimentation tank and sludge thickening tank follows this type. Zone settling can be explained by the solids-flux theory (Dick and Ewing, 1967). But this theory has some constraints blocking its application, mainly the difficulty in describing the relationship between initial settling velocity (v_i) and initial suspended solids concentration (C_i). Several types of empirical equations were developed to define an approximate relationship (Giokas et al., 2003). Vesilind's equation (1968) is one of the most used. It can be expressed as:

$$v_i = v_0 e^{-nC_i} \tag{5-1}$$

where v_0 , V_0 , n , and k are the empirical settling parameters.

4) Compression: At the bottom of secondary sedimentation tank, the particles are so close to each other that they are actually in contact. Settling can only occur by shrink of the matrix from diminishing of void spaces and squeezing out of water. This type of settling is especially important in sludge thickening tank.

Zone settling and compression settling could be the main settling processes for ASBR digestate in this study, as the SS concentrations were much higher than 5000 mg/L. Their settling behaviors can be described as Figure 5.2. As particles settle, clear water zone and compression zone expanded. At the end of zone settling process, only these two parts exist and settling occurs only from compression.

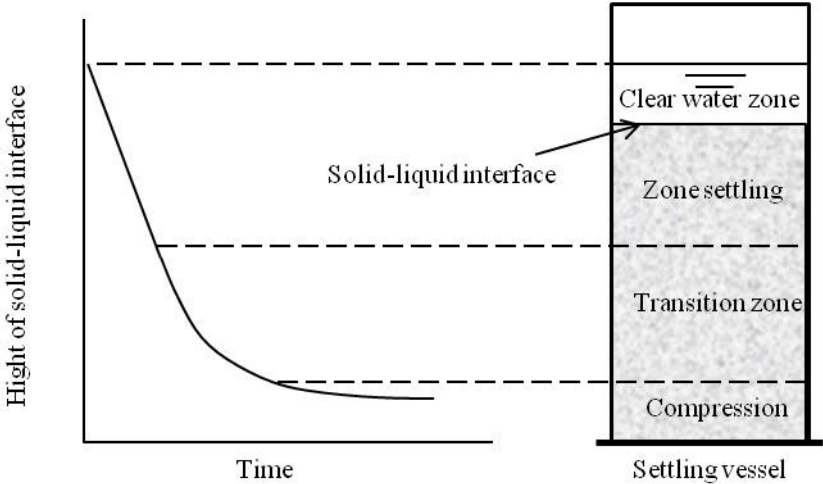


Figure 5.1 Settling of a suspension exhibiting zone settling behavior

5.3.2 Solid settling and accumulation at relative lower OLRs

For digestate from operation A-1, A-2 and A-3, obvious solid-liquid interface can be observed in 500 ml measuring vessels, as shown in Figure 5.2. Solid settlement curve was then plotted from the height of solid-liquid interface vs. time. Figure 5.3 gives some of the settlement curves from the experiment results.

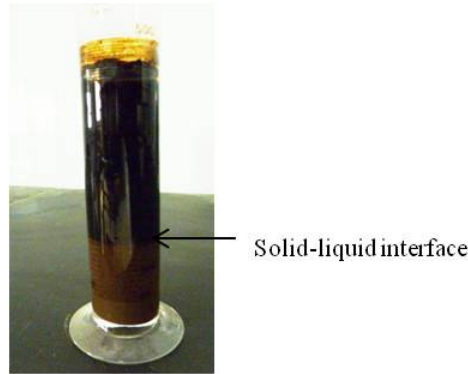


Figure 5.2 Solid-liquid interface observed

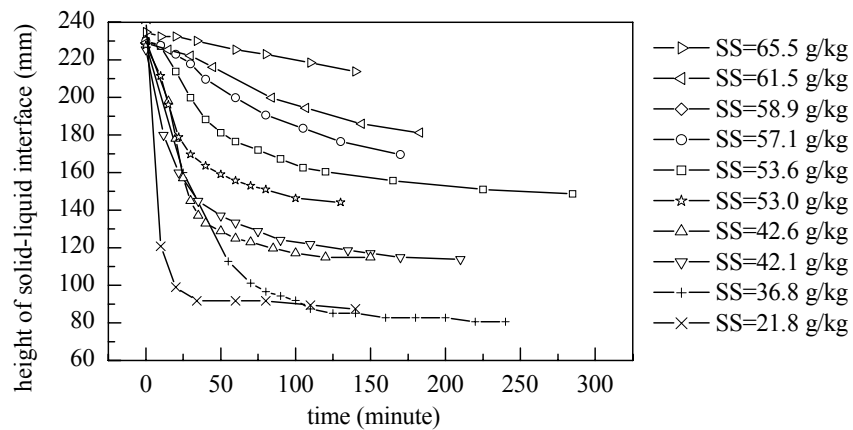


Figure 5.3 Solid settlement curves

Initial settling velocities were calculated using the linear part of the settlement curves. Their relationship with the SS concentration was then described by Vesilind equation, as shown in Figure 5.4.

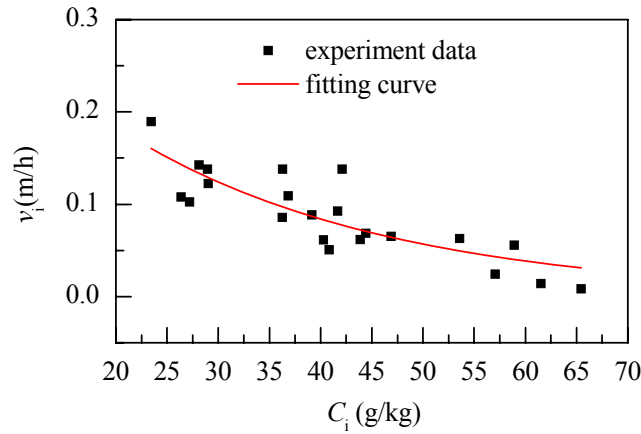


Figure 5.4 v_i - C_i relationship

Both nonlinear and linear fitting were carried out by Origin 8.0. The results were listed in Table 5.2. Results from these two fitting method were similar. As a result, only linear fitting results were used for further discussion as simplification.

Table 5.2 Simulation results from settling velocity- SS concentration data

	expression	v_0	n	R^2
nonlinear fitting	$v_i = 0.40 \times \exp(-0.04 \times C_i)$	0.40 ± 0.09	0.039 ± 0.007	0.6971
linear fitting	$\ln v_i = -0.37 - 0.06 \times C_i$	0.69 ± 0.25	0.055 ± 0.007	0.7289

From these simulation results, prediction can be made for the sludge discharging point. For example, in operation A-1, the settling time was 12.5 hours. As the feeding flux was 5L per cycle and the diameter of the digester was 0.5 m, a settling height of about 0.025 m was required. Consequently, settling velocity should be higher than 0.002 m/h.

Some factors like biogas production may interrupt the solid settling performance. To eliminate their impact, a safety factor of 5 was set according to experimental data. Settling velocity should be 5 times of the above values, which is 0.01 m/h in this case. According to the simulation result, SS concentration can reach to 74.5 g/kg for efficient settling.

For operation A-2 and A-4, as the feeding flux increased to 10L per cycle, a settling height of about 0.05 m was required. To avoid washout of solid substrates and microorganisms, SS concentrations should be kept below 60.5 g/kg. Similar calculation can be done to A-3 as well, resulting in a limiting SS concentration of 46.1 g/kg.

Results of solid accumulation proved validity of this prediction. Figure 5.5-5.7 gives variation of 4S in operation A-1, A-2 and A-3.

For operation A-1, operating lasted for 360 d. At day 84 and day 174, when SS concentration reached about 65.5 g/kg and 74.4 g/kg, increase of effluent concentration was found. To avoid further washout of the solids from effluent, voluntary sludge discharge was carried out. In the first time, 30 kg of digestate was discharged from the bottom of the digester; while in the second time, only 5 kg of the digestate was discharged. From day 250, obvious washout was found for the digester, with effluent concentrations upper to 55.1 g/kg. This happened at a SS concentration of about 74.6 g/kg, which was quite similar to the value of predicted in former paragraphs (74.5 g/kg). It can be found, when voluntary discharge of sludge was carried out, the effluent concentration can be kept in a stable range. However, when the sludge concentrations were not controlled, washout might happen and the effluent concentration might increase to a very high level. A long time of operation are needed to recover from this situation.

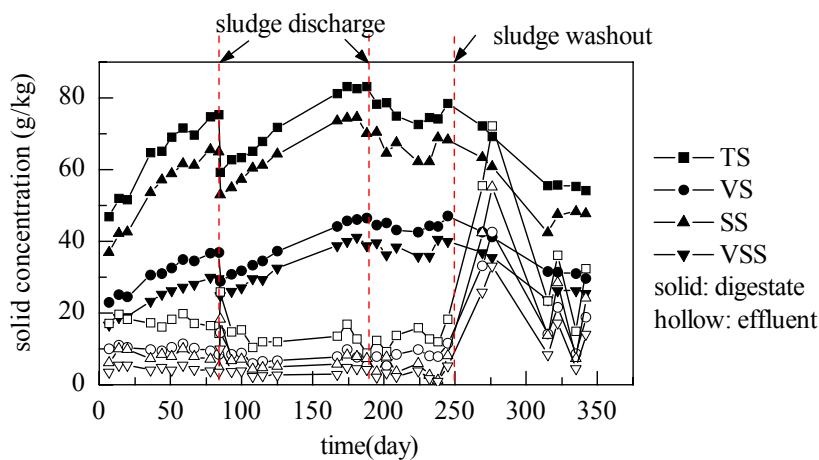


Figure 5.5 Profiles of 4S in ASBR effluent and digestate (operation A-1)

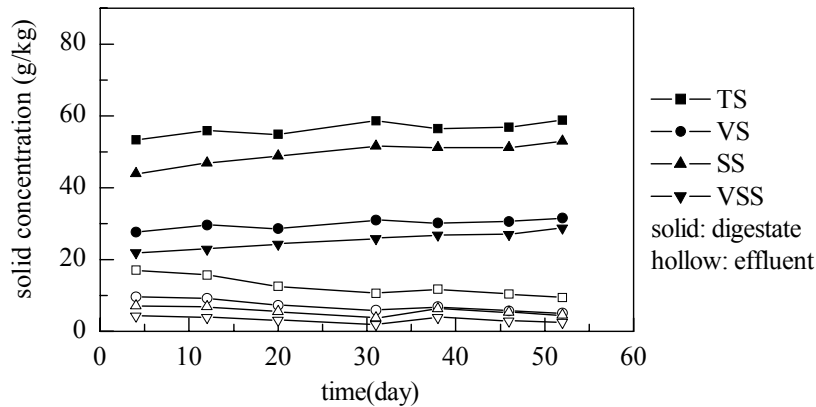


Figure 5.6 Profiles of 4S in ASBR effluent and digestate (operation A-2)

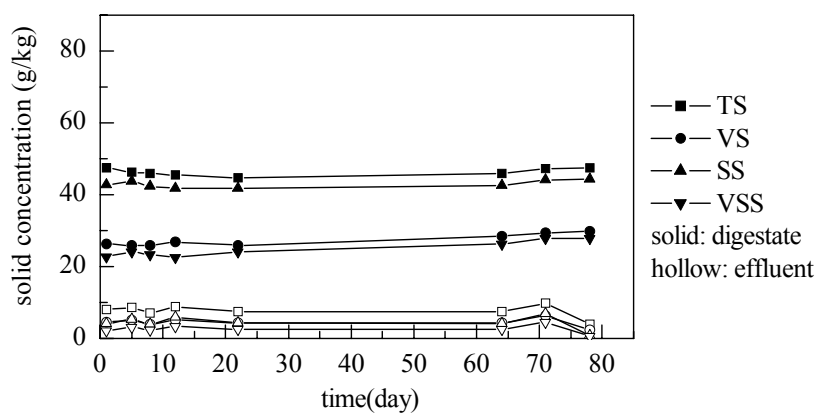


Figure 5.7 Profiles of 4S in ASBR effluent and digestate (operation A-3)

For operation A-2 and A-3, the operating time was not so long thus sludge concentration didn't reach a very high level. The highest digestate SS concentrations in operation A-2 and A-3 were 53.0 g/kg and 44.3 g/kg, respectively. Both of them were lower than the predicted values (60.5 g/kg and 46.1 g/kg) and thus no washout was found.

5.3.3 Solid settling and accumulation at a relative higher OLR: effect of biogas production

For operation A-4, the solid variations were as shown in Figure 5.8. It can be found that the effluent concentrations were much higher than those of A-2, although the requirement for settling velocity and SS concentrations were similar in these two operations. The major reason can be the higher biogas production under operation A-4. Biogas production reached to an average value of 401 l/d, while the biogas productions in operation A-1 to A-3 were less than 270 l/d. In anaerobic digestion processes, high biogas production is preferred for energy recovery. While on the other hand, biogas production may interrupt solid settlement in ASBR and thus affect the digester performance.

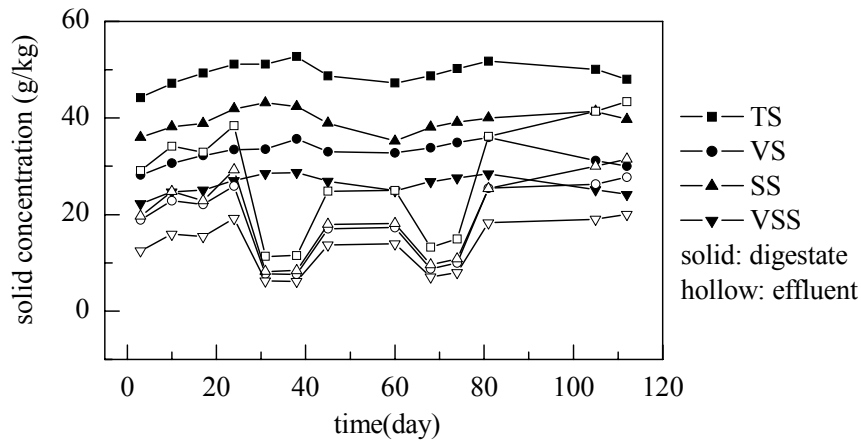


Figure 5.8 Profiles of 4S in ASBR effluent and digestate (operation A-4)

In this study, solid settlement was observed at operation A-1 to A-3. While for operation A-4, although the effluent concentrations were lower than digestate, obvious solid-liquid surface was not found due to interruption from biogas bubbles. As a result, studies were not carried out for the prediction of solid settling behaviors under this condition. Solid concentrations should be controlled according to the change of effluent quality.

Similar phenomenon has also been found in other studies. One example is the anaerobic contact process, which has a separate settlement tank for solid-liquid separation (Schroepfer et al., 1955). To remove the effect of biogas production, degasifier was used to evacuate the biogas before settlement in this process. In some studies with high production of biogas, solid-liquid separation was realized by floatation (Hur et al., 1999). Under this case, solid concentration was found highest at the top of the digester while effluent was discharged from the bottom. Although solid settlement was not realized, SRT can still be prolonged.

5.3.4 Relationship between viscosity and SS concentration

From the above discussion, it was found SS concentration affected settling performance. SS concentration should be kept below a suitable range to avoid washout of solid substrates and microorganisms. But the monitoring of SS is not so convenient that its application on ASBR control was limited. The standard method for SS determination needs more than 24 hours. Some online monitors have been developed for determination of SS (WTW, 2012). However, these equipments were expensive thus it's not available for most lab or pilot studies. As a result, a simple method was suggested for estimation of SS concentration. Relationship

between apparent viscosity and SS concentration was found, thus viscosity was used for estimation of SS. As the apparent viscosity can be quickly and easily detected by a rotational viscometer, it's easy to realize monitoring and control of the SS concentration.

Relationship between solid concentration and viscosity can be expressed in many ways, including Einstein equation (equation (5-2)) (Mardles, 1940), Brinkman equation (equation (5-3)) (Brinkman, 1952), Vand equation (equation (5-4)) (Vand, 1948) and so on.

$$\mu = \mu_0 \cdot (1 + 2.5 \times \varphi) \quad (5-2)$$

$$\mu = \mu_0 \cdot (1 - \varphi)^{-2.5} \quad (5-3)$$

$$\mu = \mu_0 \cdot \exp\left[\frac{(2.5 \times \varphi + 2.7 \times \varphi^2)}{(1 - 0.609 \times \varphi)}\right] \quad (5-4)$$

where μ is the apparent viscosity, mPa·s; μ_0 is the apparent viscosity of the medium, mPa·s; φ is volume fraction of the suspended solids, %.

For calculation, φ is needed. But this parameter is hard to be determined during operation of digesters. As a result, an exponential model was widely used instead of these equations in the biological processes (Cheng and Law, 2003; Pevere et al., 2005). SS concentration, which was much easier to be determined in these processes, was used (equation (5-5)).

$$\mu = \mu_0 \cdot \alpha \cdot e^{\beta \cdot SS} \quad (5-5)$$

which can be expressed as SS concentration in the unit of %; α is temperature-related coefficient; β is a constant.

Figure 5.9 gives viscosity and SS concentration results of operation A-1, A-2, and A-3.

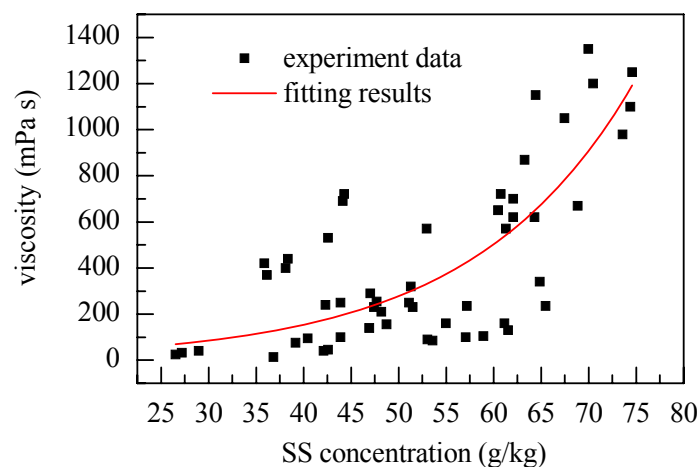


Figure 5.9 Relationship between viscosity and SS concentration

The fitting results were: $\mu_0 \cdot \alpha = 14.4 \pm 7.6$, $\beta = 0.06 \pm 0.01$, $R^2 = 0.5645$.

Combined with the results in Section 5.3.3, viscosity can be used for prediction of the sludge discharging point. As discussed, SS concentration should be below 74.5, 60.5 and 46.1 g/kg for operation A-1, A-2 and A-3. It can be calculated that the apparent viscosity should not exceed 1184 mPa·s, 518 mPa·s, and 220 mPa·s accordingly. Highest viscosity found in operation A-1, A-2 and A-3 was 1250 mPa·s, 570 mPa·s, 720 mPa·s. Simulation results were fit for those from operation A-1 and A-2. But for operation A-3, actual viscosity exceeded the predicted value while no washout happened. It probably resulted from over estimation of the safety factor.

In former calculation, safety factor (k) was used as the following way:

$$v_{required} = v_{needed} \cdot k \quad (5-3)$$

It caused over estimation as the insurance settling height increased by multiple when HRT decreased. To correct this problem, following method was used for safety insurance.

$$v_{required} = v_{needed} + v_{safety} \quad (5-4)$$

v_{safety} was introduced as a constant, which was deduced from the simulation result of operation A-1 as 0.008 m/h.

Based on equation (5-4), it was calculated that the SS concentration should not exceed 74.5, 69.5 and 61.1 g/kg. Apparent viscosity should not exceed 1185, 882 and 536 mPa·s accordingly. These results were fitter for the experiment data. However, longer operation of A-2 and A-3 are needed for further validation.

5.4 Digester stability and performance: comparison with CSTR

5.4.1 Digester stability

ASBRs were compared with CSTRs by their stability and performance under stable phases. The results were listed in Table 5.3. For operation A-1, data before washout were used. Both ASBR and CSTR were operated stably under these conditions. No significant accumulation of VFAs was found. Values of pH and VFA/alkalinity ratio were also kept inside the suitable ranges.

Table 5.3 Digester stability and performance: comparison between ASBR and CSTR

	A-1	A-2	A-3	A-4	C-1	C-2	C-3	C-4
<i>Digester stability</i>								
pH	7.42± 0.10	7.26± 0.04	7.00±0.08	7.41± 0.04	7.48± 0.11	7.31± 0.03	6.96±0.10	7.55±0.04
alkalinity (mg CaCO ₃ /L)	12037±1182	7509± 996	5554± 907	10504±1428	6349±852	6244± 831	3578±202	8065±954
VFAs (mg COD/L)	61±16	115±33	65±16	99±13	120±19	123±36	96±20	159±56
VFA/alkalinity	0.005	0.015	0.012	0.009	0.019	0.02	0.027	0.02
propionic acid in VFAs(%)	19.6	14.4	17.5	28.2	30.4	48.2	50.9	30.3
<i>TS removal</i>								
TS in feed (g/kg)	79.2 ±11.1	39.6 ±5.9	19.8 ±3.0	79.2 ±11.1	79.2 ±11.1	39.6 ±5.9	14.7 ±2.0	65.5 ± 9.5
TS in effluent (g/kg)	17.7 ± 1.3	10.9 ± 1.2	7.9 ± 1.9	27.4±11.6	28.6 ± 3.9	25.8 ± 1.0	10.9 ± 0.3	36.7 ± 5.3
TS accumulated (g/(kg·d))	0.36	0.17	0.10	0.23	0	0	0	0
TS removal rate (%)	77.6	72.4	59.9	65.4	63.8	35.0	25.7	44.0
TS degradation rate (%)	62.0	65.2	55.5	60.5	63.8	35.0	25.7	44.0
<i>VS removal</i>								
VS in feed (g/kg)	61.7±9.1	30.8 ±1.6	15.4 ±0.8	61.7 ± 9.1	61.7 ± 9.1	30.8 ±1.6	11.6 ±1.8	50.2 ± 11.2
VS in effluent (g/kg)	10.2±0.6	6.2±0.9	4.5±1.2	18.3±7.6	19.3 ± 1.8	15.0 ± 0.2	6.6 ± 0.2	21.6 ± 4.5
VS accumulated (g/(kg·d))	0.19	0.14	0.07	0.19	0	0	0	0
VS removal rate (%)	83.4	80.0	70.6	70.4	68.7	51.5	43.1	57.0

Table 5.3 (-continued)

	A-1	A-2	A-3	A-4	C-1	C-2	C-3	C-4
VS degradation rate (%)	73.2	72.6	66.7	65.1	68.7	51.5	43.1	57.0
<i>SS removal</i>								
SS in feed (g/kg)	42.5 ± 6.3	21.2 ± 3.2	10.6 ± 0.6	32.4 ± 1.9	42.5 ± 6.3	21.2 ± 9.1	7.0 ± 0.6	32.4 ± 1.9
SS in effluent (g/kg)	8.2 ± 2.8	5.1 ± 0.9	4.6 ± 1.9	19.7 ± 8.4	18.2 ± 1.4	14.1 ± 1.0	6.5 ± 1.9	25.0 ± 8.4
SS accumulated (g/(kg·d))	0.34	0.17	0.09	0.21	0.34	0.17	0.09	0.21
SS removal rate (%)	80.6	76.1	56.5	53.5	57.1	33.6	6.0	23.0
SS degradation rate (%)	53.5	62.6	49.0	45.3	57.1	33.6	6.0	23.0
<i>VSS removal</i>								
VSS in feed (g/kg)	31.1 ± 6.3	15.5 ± 4.2	7.8 ± 2.1	31.1 ± 6.3	31.1 ± 6.3	15.5 ± 4.2	5.3 ± 0.5	22.7 ± 1.8
VSS in effluent (g/kg)	4.6 ± 0.7	3.3 ± 0.9	2.6 ± 1.2	13.5 ± 5.2	14.9 ± 1.6	11.6 ± 0.7	4.4 ± 1.3	15.2 ± 2.2
VSS accumulated (g/(kg·d))	0.18	0.14	0.06	0.19	0	0	0	0
VSS removal rate (%)	85.3	79.0	66.3	56.5	52.0	25.3	17.0	33.1
VSS degradation rate (%)	66.0	64.1	59.3	46.3	52.0	25.3	17.0	33.1

Table 5.3 (-continued)

	A-1	A-2	A-3	A-4	C-1	C-2	C-3	C-4
<i>Biogas production</i>								
biogas production(l/d)	262±17	265±21	233±11	494±18	260±38	225±32	150±18	476 ± 91
specific biogas production (l/(l·d))	1.53 ± 0.10	1.47 ± 0.14	1.23 ± 0.06	2.74 ± 0.10	1.48± 0.22	1.25±0.18	0.70±0.11	2.6 ± 0.5
biogas yield (l/g VS _{added})	0.87 ± 0.06	0.86 ± 0.07	0.76± 0.04	0.80± 0.03	0.84 ± 0.12	0.73± 0.10	0.66±0.17	0.77±0.15
methane content (%)	69.1 ± 0.5	69.3 ± 1.2	72.6 ± 0.7	70.7 ± 2.4	69.2 ± 0.3	68.8 ± 0.4	68.2 ± 0.1	69.6 ± 0.0
methane production(l/(l·d))	1.05 ± 0.07	1.02 ± 0.08	0.89 ± 0.04	1.94 ± 0.07	1.02 ± 0.14	0.85 ± 0.12	0.54±0.07	1.82 ± 0.06
methane yield (l/g VS _{added})*	0.53 ± 0.03	0.53 ± 0.04	0.49 ± 0.03	0.50 ± 0.02	0.52 ± 0.08	0.45 ± 0.06	0.39±0.10	0.48 ± 0.11

*methane yield under a standard state of 273.15K, 101.325KPa.

However, higher values of alkalinity were found in ASBR, which resulted in lower VFA/alkalinity ratio. It implied better buffering ability to feeding shock in ASBR. The proportion of propionic acid in VFAs was also lower in ASBR.

These results probably resulted from accumulation of solids inside the ASBRs. Average TS concentrations were 67.4 g/kg, 55.9 g/kg, 51.7 g/kg, and 50.0 g/kg in A-1, A-2, A-3 and A-4. While in CSTR, average TS concentrations were only 28.6 g/kg, 25.8 g/kg, 10.9 g/kg, and 27.7 g/kg for C-1, C-2, C-3 and C-4. By solid settling, more buffering materials were kept inside ASBR and thus improved digester stability. However, the solid concentration should be controlled to avoid solid washout.

5.4.2 Digester performance

ASBR was compared with CSTR by substrates removal and biogas production. Results were also listed in Table 5.3.

For ASBR, as the effluent concentrations were lower than those of digestate, both apparent removal rate and degradation rate were calculated. Apparent removal rate was calculated from data of effluent as following:

$$\frac{C_{\text{feed}} - C_{\text{effluent}}}{C_{\text{feed}}} \times 100 \quad (5-5)$$

where C means the concentration of specific indicators, e.g.: 4S, COD.

Degradation rate means the ratio of substrates that were actually biodegraded. It can be calculated from the following equation:

$$\frac{C_{\text{feed}} - C_{\text{effluent}} - C_{\text{accumulated}}}{C_{\text{feed}}} \times 100 \quad (5-6)$$

For CSTR, no accumulation happened when digesters reached a stable phase. The apparent removal rate equals to the degradation rate.

ASBR was found outperforming CSTR in many aspects.

Firstly, the effluent quality was better in ASBR. Despite of high solid concentrations inside the digester, ASBR effluent had lower solid concentrations than CSTR due to solid settlement. TS concentrations in ASBR effluent were 17.7 g/kg, 10.9 g/kg, 7.9 g/kg, and 27.4

g/kg for operation A-1 to A-4, compared with 28.6 g/kg, 25.8 g/kg, 10.9 g/kg, and 36.7 g/kg for C-1 to C-4. It implied less cost was needed for further disposal of the effluent.

Secondly, with solids settled inside the digester, ASBR also performed better in degradation of organic substrates, especially suspended organics. The VSS removal rate was 14.0%, 38.8%, 41.9%, and 13.2% higher than that of CSTR in operation 1-4. Accordingly, more biogas was formed in ASBR. Specific methane yield was 0.49-0.53 l/g VS_{added}, compared with 0.39 - 0.52 l/g VS_{added} in CSTR. With more methane produced, ASBR presented better ability in energy recovery. The detailed calculation about energy balance will be carried out in later chapters.

Moreover, it can be found effect of HRT to digester performance was more obvious in CSTR. VS degradation rate in CSTR decreased from 68.7% to 43.1% when HRT decreased from 35 d to 9.5 d. While for ASBR, only a slight decrease from 73.2% to 66.7% was found. Difference was more significant in VSS degradation rate. For ASBR, it decreased from 66.0% to 59.3%; while for CSTR, it crashed from 52.0% to 17.4%. It proved more suspended substrates were degraded when ASBR was applied.

It implied that HRT can be lower for ASBR to reach a certain digester performance. For example, to achieve a VS degradation rate of 50% in this study, HRT of the CSTR digester should be kept above 18 d. While for ASBR, HRT can be lower than 9.5 d. This result was especially meaningful in practical application. As lower HRTs can be chosen, volume of the digester can be reduced and also the capital cost for infrastructure construction.

Former lab-scale studies (Hou et al., 2010) also proved similar advantage of ASBR over CSTR. As shorter settling height was needed in lab digesters, the solid-liquid separation was even better in the lab digesters. Under a HRT of 20 d and an OLR of 2 kg VS/m³/d (which was quite similar to the condition of operation A-1), SRT/HRT ratio can be over 10. VS removal rate reached 88.6%, compared with 61.2% by CSTR under the same operating condition.

5.5 Conclusion

In this chapter, ASBR was applied for anaerobic digestion of thMBW. It was found:

1) Solid settlement was realized in ASBR. At lower OLRs of 1.6-1.8 kg VS /($m^3 \cdot d$), obvious solid-liquid interface can be found. Settling behaviors can be described by combination of zone settling compression settling, and expressed by the Vesilind equation: $v_i = 0.69 \times \exp(-0.06 \times SS_i)$. However, at a higher OLR of 3.4 kg VS /($m^3 \cdot d$), settling performance was interrupted by biogas production and could not be explained by the classical theory.

2) Viscosity was found related to the SS concentration. Both of them can be used for prediction of the washout point when settling is not interrupted by biogas production. It was predicted that, SS concentrations should be kept below 74.5 g/kg, 60.5 g/kg, and 46.1 g/kg for operation A-1, A-2, and A-3, respectively. Accordingly, viscosity should be kept below 1185 mPa·s, 518 mPa·s, and 220 mPa·s for these operations. Experiment data were found corresponding with these prediction results.

3) With higher alkalinity and less accumulation of propionic acid, ASBR showed better stability than CSTR.

4) ASBR outperformed CSTR in many aspects, including lower effluent strength, better removal of organics, and more biogas production. Especially, ASBR performance showed better resistance to HRT decrease than CSTR, which indicated digester volume can be reduced when ASBR is applied.

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Chapter 6 Modeling of process kinetics: using SRT as an important parameter

6.1 Introduction

Anaerobic digestion is a complicated process which has complex substrates, multiple microbial populations, and sophisticated reactions. At present, the most comprehensive description of this process is the ADM1 model developed by International Water Association (Batstone et al., 2002). However, this model is too complicated to be solved and is limited to modeling of CSTR. Some researchers developed simplified models for description of the anaerobic processes in ASBR (Bagley and Brodkorb, 1999; Zheng, 2003), but most of them were used for ASBR treating sewage and low strength wastewater.

ASBR outperforms CSTR due to prolonging the SRT without changing HRT. As a result, SRT must be a very important parameter in the description of the process kinetics in ASBR.

SRT usually means the average time that biosolids (microorganisms) are in the reactor. It is an important parameter in design and operation of biological reactors. When SRT decreases, microorganisms' loss rate exceeds their growth rate and thus "washout" happens. As a result, SRT should be kept above a certain value for stable operation of the reactors. In activated sludge system, typical SRTs are 4-10 d. While in anaerobic digestion processes, as the growth rates of methanogens are much slower (about 0.28 d^{-1} for acetoclastic methanogens), longer SRTs are needed. SRTs are suggested for at least 10 d and the typical range is 15-25 d (Rittmann and McCarty, 2001).

For anaerobic digestion of high-solids-content waste, definition of SRT may be broader. Distinguished from wastewater or low-solids-contents waste digestion, solids in the digestate include not only biosolids, but also organic solid substrates. As a result, the concepts of SRT_X (SRT of microorganisms) and SRT_P (SRT of organic particulate substrates) are needed for discussion under these cases. In anaerobic digestion of simple organics like sugar and protein, methanogenesis is the limiting step. Consequently, it's important to keep SRT_X in proper ranges. While for digestion of complex high-solids-content waste, solid hydrolysis is the

limiting step. Under this condition, digester performance may be more dependent on SRT_P than SRT_X .

For CSTR digesters, SRT equals to HRT. While in ASBR, SRT can be separated from HRT. ASBR application to the treatment of biowaste has been carried out by some researchers. Moo Hur et al. (1999) treated high-solids-content waste by lab-scale ASBRs. Different solid-liquid separation patterns were found under mesophilic and thermophilic status. Luo et al. (2009) used ASBR for digestion of high-solids-content cassava stillage. COD removal over 80% can still be found in this study under HRT of 5 d. Lee et al. (2001) used ASBR for the digestion of manure. Biogas production in ASBR was found 205-220% of that in CSTR under HRT of 10 d.

Moreover, series studies were done by Zaiat and his fellows by adding carriers into the ASBR digesters (Siman et al., 2004; Damasceno et al., 2007; Miqueleto et al., 2005). Biofilm can be formed in the surface of carriers and thus SRT_X can be separated from SRT_P as well. Similar study was also done by Wang et al. (2009). They used anaerobic sequencing batch biofilm reactor (ASBBR) for anaerobic digestion of thermally hydrolyzed municipal biowaste. Lipid-P was used to represent the amount of microorganisms. SRT_X was calculated as 89-150 d when HRT was 13.3 d, yet apparent SRT represented in VSS was only about 24-36 d.

In this Chapter, simplified models were established to describe substrates removal of thMBW and growth of microorganisms. SRT was used as an important factor for process modeling. Especially, solid accumulation behaviors in ASBR were also qualitatively described.

6.2 Methodology

6.2.1 Rate limiting step in anaerobic digestion of thMBW

ADM 1 is a complicated model without numerical solution. By indentifying the rate limiting step, simplified models can be developed to get more concrete description for anaerobic digestion of thMBW.

ADM 1 model can be simplified to a two-step model including hydrolysis (acidogenesis) and methanogenesis. This model was firstly brought up by Buswell and Neave (1930). In the

first step, complex organics are degraded into short-chain fatty acids (VFAs) and alcohols, accompanied with production of CO₂, H₂S, H₂, and ammonia. In the second phase, methanogens utilize the intermediate products to produce CH₄ and CO₂.

From the results of Chapter 3, it was found that 52% of the organics in thMBW were composed by particulate solids, and 9% was composed by VFAs. Accordingly, the rate limiting step was assumed to be disintegration and hydrolysis in the first step (Gossett and Besler, 1982). These processes are assumed to be first order from empiric simplification (Eastman and Ferguson, 1981; Pavlostathis and Gossett, 1986) and can be expressed:

$$r_h = \frac{dP}{dt} = -k_h P \quad (6-1)$$

where P means the degradable organic particulates, g/kg; k_h means hydrolysis rate coefficient, d⁻¹.

These hydrolysis processes are mainly catalyzed by extracellular enzyme, not the microorganisms. As a result, the concentration of microorganisms has little effect on the hydrolysis rate and digestion performance.

6.2.2 Mass balance in CSTR and ASBR

As suspended solids are the main source of substrates in thMBW, it was assumed that all the substrates are comprised of particulates. As the processes are rate-limited by hydrolysis, no soluble intermediates are accumulated. Mass flow of CSTR and ASBR can then be described as Figure 6.1.

For CSTR under a steady state, the effluent properties are the same as the digestate inside the reactor. For ASBR, due to solid settlement, the concentration of particulate substrates and microorganisms are higher than effluent. Supposing all the solids have same sizes and shapes, following relationship can be introduced:

$$\frac{P}{P_e} = \frac{X}{X_e} = \frac{\theta_p}{\theta} = \frac{\theta_x}{\theta} = f \quad (6-2)$$

where f means the degree of solid separation.

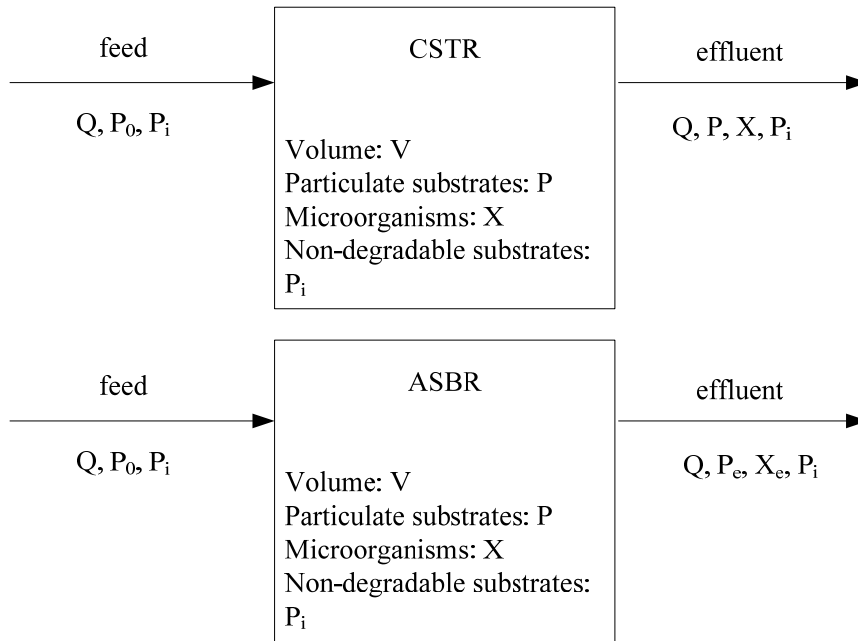


Figure 6.1 Scheme of CSTR and ASBR

As suspended solids are the main source of substrates, ASBR can be approximately simulated as a larger CSTR with a volume of $f \cdot V$ as shown in Figure 6.2. This simulated CSTR has the same HRT as the ASBR, which is $f \cdot \theta$ and equals to SRT in the ASBR. CSTR can also be expressed by this figure, taking f as 1.

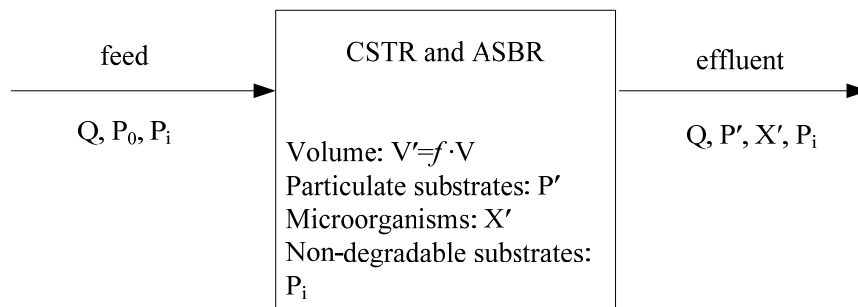


Figure 6.2 Scheme of ASBR as an approximation of CSTR

6.2.3 Effect of SRT on digestion efficiency: from the view of hydrolysis limitation

Following mass balance can be made for the degradable particulate substrates in a steady CSTR:

$$(P_0 - P) \cdot Q = V' \cdot (-r_h) \quad (6-3)$$

where P_0 is the degradable particulates in feed, g/kg; Q means the flow rate, m³/d; V' means the simulated digester volume, m³;

As discussed before,

$$\frac{V'}{Q} = f \cdot \frac{V}{Q} = f \cdot \theta = SRT \quad (6-4)$$

Combining equation (6-1), (6-3) and (6-4), the degradable particulates concentration can be expressed as:

$$P = \frac{P_0}{1 + SRT \cdot k_h} \quad (6-5)$$

Similar mass balance can be found for the microorganisms:

$$(X_0 - X) \cdot Q = V \cdot (-r_x) \quad (6-6)$$

where X_0 means the amount of microorganisms in the feed, which equals to 0 in this study; X means the amount of microorganisms in the digester.

Microorganisms' growth rate can be related to the consumption rate of substrates, which can be expressed as (Rittmann and McCarty, 2001):

$$r_x = \frac{dX}{dt} = Y \left(-\frac{dP}{dt} \right) - bX \quad (6-7)$$

where Y means the specific yield of microorganisms as a whole to the substrate, g_X/g_P ; b means the decay rate of microorganisms, d^{-1} .

Combining equation (6-1), (6-3), (6-6) and (6-7), X can be calculated out:

$$X = Y \cdot \frac{P_0 - P}{1 + b \cdot \theta} \quad (6-8)$$

VSS can be considered as the mixture of both the microorganisms and particulate substrates, including non-degradable substrates as well:

$$VSS = X + P + P_i \quad (6-9)$$

$$VSS_0 = P_0 + P_i \quad (6-10)$$

Degradable rate of VSS (R_{VSS}) can be derived from VSS and VSS_0 as:

$$R_{VSS} = \frac{VSS_0 - VSS}{VSS_0} \quad (6-11)$$

Accordingly, degradation rate of VSS (R_{VSS}) can be expressed by the following equation:

$$R_{vss} = \alpha \cdot \left(1 - \frac{1}{1 + k_h \cdot SRT}\right) \cdot \left(1 - \frac{Y}{1 + b \cdot SRT}\right) \quad (6-12)$$

where α means the proportion of degradable substrates in feed, which equals to P_0/VSS_0 .

In anaerobic digestion processes, the values of Y are usually lower than 1. Rittmann and McCarty (2001) suggested the value of Y for acetoclastic methanogens is about 0.04 g/gAc and for hydrogenotrophic methanogens it is 0.45 g/gH₂. Pavlostathis and Gossett (1986) found a Y value of 0.057 for methanogenesis and 0.2 for hydrolysis in the anaerobic digestion of WSS. For the parameter b , its value has little difference for various microorganisms. Rittmann and McCarty (2001) suggested b as 0.03 d⁻¹ for both acetoclastic and hydrogenotrophic methanogens. Pavlostathis and Gossett (1986) found a b value of 0.015 for methanogenesis and 0.1 for hydrolysis. In ADM1 model, suggested values for Y and b are 0.04-0.1gCOD/gCOD and 0.02 d⁻¹.

According to these, value of Y and b was set as 0.1 g/g and 0.02 d⁻¹. Under this condition, the third part of equation (6-12) was quite close to 1, especially for ASBR with a long SRT. If ignoring this part, equation (6-12) turns to:

$$R_{vss} = \alpha \cdot \left(1 - \frac{1}{1 + k_h \cdot SRT}\right) \quad (6-13)$$

6.2.4 Effect of SRT on digestion efficiency: using a model from Romero (1991)

Romero (1991) established a model for simulation of the anaerobic digestion processes treating municipal biowaste. His model was proved fit with the results of several studies (Pérez, 2001a; Pérez, 2001b; De la Rubia, 2006; Fernández, 2010; Fdez.-Güelfo, 2011) and can be expressed as:

$$-r_P = \mu_{\max} \frac{(P_0 - P) \cdot P}{P_0} \quad (6-14)$$

where P was not limited to particulates but all the organic substrates.

By including a mass balance equation, it was found:

$$P = \frac{P_0}{SRT \cdot \mu_{\max}} \quad (6-15)$$

$$X = \frac{Y}{1+b \cdot SRT} \cdot (P_0 - P) \quad (6-16)$$

Degradation rate of VSS can be described as:

$$R_{VSS} = \alpha \cdot \left(1 - \frac{Y}{1+b \cdot SRT}\right) \cdot \left(1 - \frac{1}{SRT \cdot \mu_{max}}\right) \doteq \alpha \cdot \left(1 - \frac{1}{SRT \cdot \mu_{max}}\right) \quad (6-17)$$

Equation (6-13) and (6-17) are also used for description of the degradation rate of SS and VS. However, some parts of the substrates, like inorganic substrates and soluble organics were ignored for simplification.

6.2.5 Prediction of effluent in CSTR and ASBR

From the above discussion, effluent concentration can be expressed as a function of SRT as well.

For digesters, following mass balance can be made:

$$Q \cdot C_0 - Q \cdot C_e - V \cdot R_{acc} = Q \cdot C_0 \cdot R \quad (6-18)$$

where C_e means the effluent concentration of any indicator under this simulation scenery, e.g.: VSS, SS, or VS, g/kg; R means the removal rates of these indicators; R_{acc} means the accumulation rate of these substrates, g/(kg·d).

$$C_e = C_0 \cdot (1-R) - \theta \cdot R_{acc} \quad (6-19)$$

From equation (6-13), effluent concentration can be calculated as:

$$C_e = C_0 - C_0 \cdot \alpha \cdot \left(1 - \frac{1}{1+k_h \cdot SRT}\right) - \theta \cdot R_{acc} \quad (6-20)$$

From equation (6-17) introduced from Romero's model, effluent concentration can be expressed as:

$$C_e = C_0 - C_0 \cdot \alpha \cdot \left(1 - \frac{1}{SRT \cdot \mu_{max}}\right) - \theta \cdot R_{acc} \quad (6-21)$$

For CSTR under a steady state, R_{acc} equals to 0.

6.2.6 Prediction of solid accumulation in ASBR

During practical operation of ASBR, prediction of solid accumulation behaviors is important. As discussed in Chapter 5, solid settling velocity was affected by solid

concentration and sludge washout might happen when solid concentration reached a relative high level.

When effluent concentrations were known, prediction of solid accumulation can be made based on equation (6-20) or (6-21). Solid accumulation rate can be calculated as:

$$R_{acc} = \frac{1}{\theta} \cdot \left[C_0 - C_0 \cdot \alpha \cdot \left(1 - \frac{1}{1 + k_h \cdot SRT} \right) - C_e \right] \quad (6-22)$$

Or following equation based on Romero's model.

$$R_{acc} = \frac{1}{\theta} \cdot \left[C_0 - C_0 \cdot \alpha \cdot \left(1 - \frac{1}{SRT \cdot \mu_{max}} \right) - C_e \right] \quad (6-23)$$

6.3 Data source

Data listed in Table 5.3 was used for model simulation. To improve reliability of the simulation, results from two lab scale operations of CSTR (C-5, C-6) and one pilot operation of ASBR (A-5) were also used.

Lab CSTR was as shown in Figure 6.3. It was made by glass, with a total volume of 30 L and effective volume of 25L. Mechanical agitator was used for mixing, with agitation rate of 100rpm and strength of 5 minutes/ hour. The digesters were kept in mesophilic phase by water bath, with temperature range from 35°C to 37°C. As same as the pilot digester, semi-continuous operation mode was used. Daily feeding and discharging were carried out. Feed materials were diluted by tap water to meet the requirement of OLR and HRT.

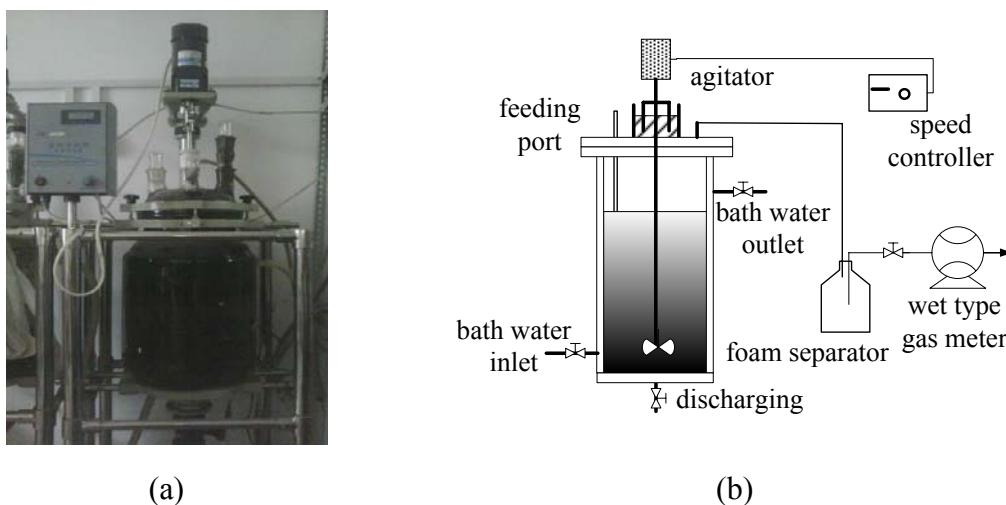


Figure 6.3 lab digester: (a) photo image; (b) sketch

Operating parameters and digester performance of these 3 operations were listed in Table 6.1.

Table 6.1 Operating parameters and results of operation C-5, C-6, and A-5

	C-5	C-6	A-5
<i>operating parameters</i>			
Digester scale	lab	lab	pilot
HRT (d)	10	5	12.3
OLR (kg VS/(m ³ ·d))	1.2 ± 0.7	1.2 ± 0.7	5.0 ± 0.6
Operating time (d)	20	15	71
<i>stability</i>			
pH	7.12 ± 0.04	6.94 ± 0.16	7.54 ± 0.05
alkalinity (mg CaCO ₃ /L)	3173 ± 280	1958 ± 100	8368 ± 639
VFAs (mg COD/L)	138 ± 20	814 ± 107	130 ± 19
VFA/alkalinity	0.043	0.416	0.016
propionic acid in VFAs(%)	39.3	66.2	16
<i>4S removal</i>			
TS in feed (g/kg)	15.2 ± 0.8	7.6 ± 0.4	79.2 ± 11.1
TS in effluent (g/kg)	11.6 ± 1.5	7.1 ± 1.1	34.2 ± 8.6
TS degradation rate (%) ¹	23.7	6.5	51.7 (56.9, 0.36)
VS in feed (g/kg)	11.9 ± 0.7	5.9 ± 0.3	61.7 ± 9.1
VS in effluent (g/kg)	6.9 ± 0.8	4.6 ± 0.2	20.0 ± 4.8
VS degradation rate (%) ¹	42.1	21.8	64.3 (67.5, 0.17)
SS in feed (g/kg)	7.5 ± 0.7	3.8 ± 0.3	42.5 ± 6.3
SS in effluent (g/kg)	6.8 ± 1.5	3.6 ± 1.0	25.6 ± 9.3
SS degradation rate (%) ¹	9.5	4.7	30.0 (39.7, 0.36)
VSS in feed (g/kg)	5.7 ± 0.5	2.8 ± 0.3	31.1 ± 6.3
VSS in effluent (g/kg)	4.8 ± 0.9	2.7 ± 0.6	15.2 ± 5.1
VSS degradation rate (%) ¹	15.7	6.3	44.8 (51.1, 0.17)

Table 6.1 (-continued)

	C-5	C-6	A-5
<i>biogas production</i>			
biogas production(l/d)	16.3±0.8	11.0±0.5	669±82
specific biogas production (l/(l·d))	0.65 ± 0.03	0.44 ± 0.02	3.62 ± 0.44
biogas yield (l/g VS _{added})	0.55 ± 0.03	0.37 ± 0.02	0.72± 0.09
methane content (%)	73.6 ± 0.1	69.8 ± 0.3	71.4 ± 1.5
methane production(l/(l·d))	0.48 ± 0.02	0.31 ± 0.01	2.58 ± 0.32
methane yield (l/g VS _{added}) ²	0.35 ± 0.02	0.23 ± 0.01	0.46 ± 0.08

¹For ASBR, the result was expressed as: degradation rate (apparent removal rate, accumulation rate)

²Methane volume was adjusted to that under standard conditions: 101.325 kPa and 273.15 K.

For CSTR, SRT equals to HRT. For ASBR, SRT was calculated according to the following equation:

$$SRT_i = HRT \cdot \frac{C_{i, \text{digestate}}}{C_{i, \text{effluent}}} \quad (6-24)$$

where $C_{i, \text{digestate}}$ is the concentration of substrates in digestate, which was expressed a specific indicator I ; $C_{i, \text{effluent}}$ means the concentration of substrates in effluent;

For example, SRT_{TS} means the retention time of total solids and is calculated from TS concentrations in the digestate and effluent. Results were listed in Table 2.

Table 6.2 SRTs of ASBRs

	A-1	A-2	A-3	A-4	A-5
HRT (d)	35	18	9.5	18	12.3
SRT_{TS} (d)	138.2	92.1	55.4	30.4	18.0
SRT_{SS} (d)	256.1	175.4	88.4	34.3	18.9
SRT_{VS} (d)	115.9	86.7	57.3	28.7	20.9
SRT_{VSS} (d)	207.6	156.9	90.2	30.7	21.9

6.4 Simulation results

6.4.1 Simulation of substrate degradation rate

Simulation of substrate degradation rate was carried out based on equation (6-11) and (6-16), for VSS, SS and VS, respectively. Fitting curves were as shown in Figure 6.4-6.9.

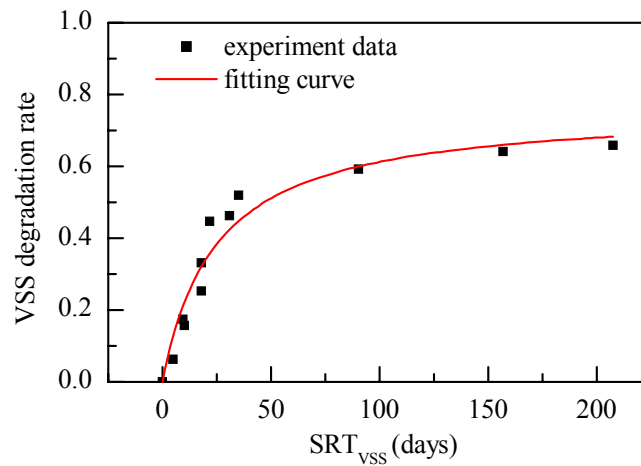


Figure 6.4 Simulation of VSS results: model based on hydrolysis limitation

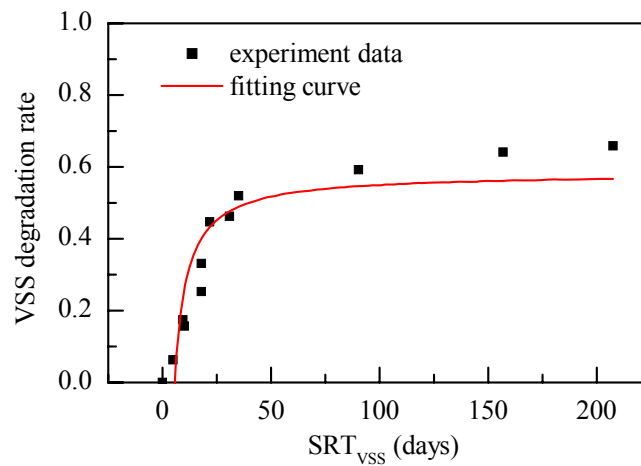


Figure 6.5 Simulation of VSS results: a model from Romero (1991)

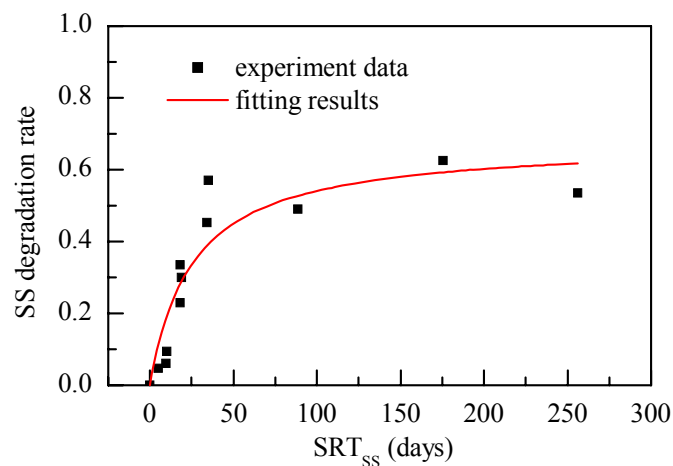


Figure 6.6 Simulation of SS results: model based on hydrolysis limitation

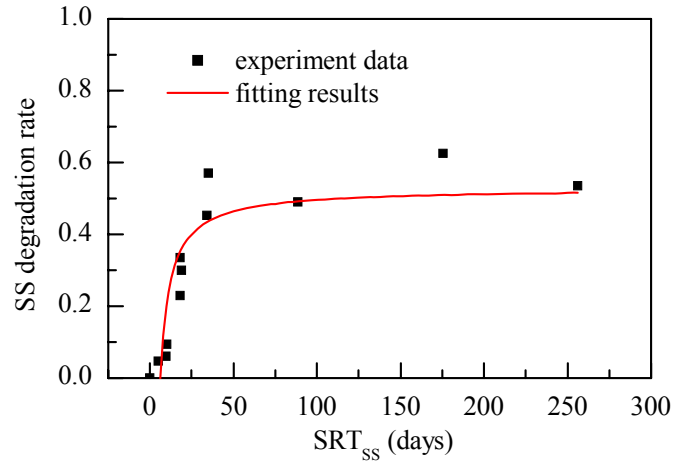


Figure 6.7 Simulation of SS results: a model from Romero (1991)

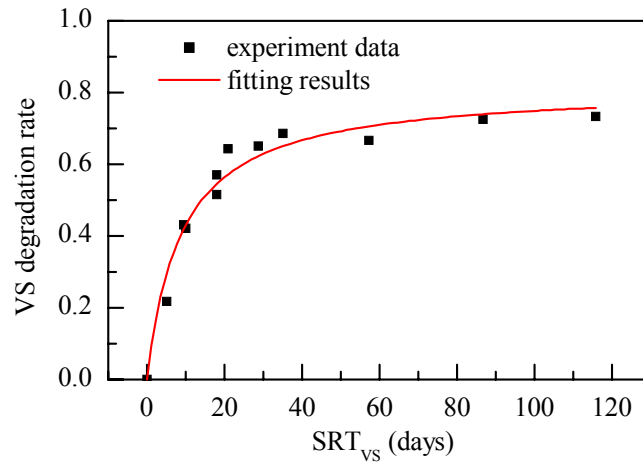


Figure 6.8 Simulation of VS results: model based on hydrolysis limitation

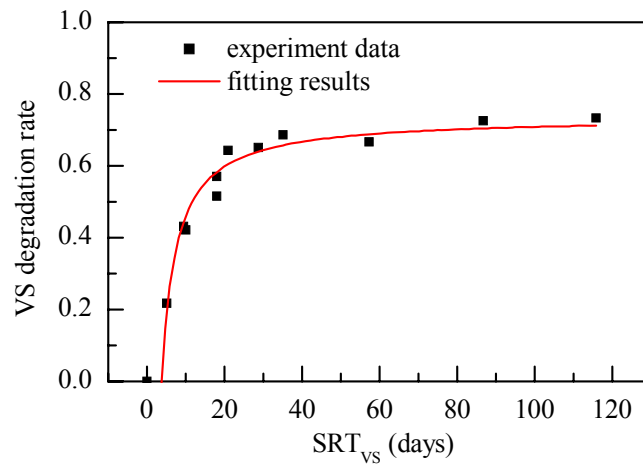


Figure 6.9 Simulation of VS results: a model from Romero (1991)

The simulation results of hydrolysis rate were listed in Table 6.3 and 6.4. It can be found simulation based both models was considered acceptable, with most R-square values higher than 0.8.

Table 6.3 Simulation results of hydrolysis rate

	α (%)	k_h	R^2	Temp.	references
VSS	76.4 ± 5.2	0.04 ± 0.01	0.9420	35-37°C	this study
SS	67.9 ± 8.2	0.04 ± 0.01	0.8416	35-37°C	this study
VS	81.6 ± 3.3	0.11 ± 0.02	0.9661	35-37°C	this study
VSS, household waste		0.1		37°C	Vavilin et al., 2005
VSS, newspaper		0.057		35°C	Vavilin et al., 2004
VSS, MBW		0.12		35°C	Liebetrau et al., 2004
VSS, fiber		0.066		35°C	Liebetrau et al., 2004

From Table 6.3, it can be found hydrolysis rate results calculated from VSS and SS were similar, while the hydrolysis rate simulated from VS was higher. This was due to the soluble organics inside VS. This part of organics contributed to the degradation rate of VS, which was ignored during simulation and counted as a part of particulates hydrolysis.

The non-degradable fraction in SS (32%) was higher than VSS (24%), which accorded with the former assumption, as SS contains non-degradable inorganic particulates as well. This fraction was lowest for VS (18%), which means the soluble organics had less non-degradable fraction than the particulate organics.

Hydrolysis rate results were compared with those from other studies in Table 6.3. It can be found the result from VS data was similar to the results of household waste and MBW, which may have similar component as this study. The results from VSS and SS data were similar to that of newspaper and fiber, which were mainly composed by solids.

The simulation results based on Romero's model were listed in Table 6.4. The degradable fractions simulated from this model were lower than that in Table 6.3, but had the same relationship among results from VSS, SS and VS (VS > VSS > SS). Simulation based on VS was similar to the values from Mata-Alvarez (1996) and Rittmann and McCarty (2001). Its

degradable fraction was similar to that from Fdez-Güelfo's study (2011), in which simulated MBW was used.

Table 6.4 Simulation results of specific microbial growth rate

microorga- nisms	substrates	α (%)	μ_{\max}	R^2	temp	references
overall	VSS	58.2 ± 3.9	0.18 ± 0.02	0.8521	35-37°C	this study
overall	SS	52.8 ± 4.5	0.17 ± 0.02	0.7898	35-37°C	this study
overall	VS	73.6 ± 1.5	0.27 ± 0.01	0.9771	35-37°C	this study
aceticlastic methanogens	VS (sewage)		0.28		35°C	Rittmann, 2001
overall	VS (simulated MBW)	73	0.58		55°C	Fdez-Güelfo, 2011
overall	VS (MBW)		0.29		35°C	Mata-Alvarez, 1996

Figure 6.4-6.9 also showed, removal of organics varied a lot when SRTs were lower than about 20 days. This value is similar to that usually suggested for stable anaerobic digestion. Digester capacity is limited for CSTR as SRTs have to be kept higher than 20 days. However, in ASBR, SRTs could get higher than 20 days when HRTs are much lower. Digester capacity can then be improved.

6.4.2 Simulation of effluent concentration

Simulation of effluent concentration was carried out based on equation (6-19) and (6-20). Results were listed in Figure 6.10.

It showed that simulation based on both the two models were acceptable, with most of the predicted results closer to the value of experimental data.

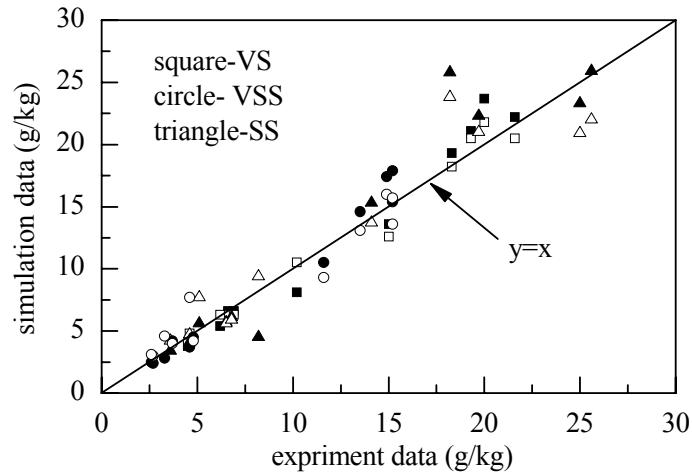


Figure 6.10 Simulation of effluent concentrations: solid- data based on the model assuming hydrolysis limitation; hollow- data based on the model from Romero (1991)

6.4.3 Prediction of solid accumulation in ASBR

Equation (6-21) and (6-22) can be used for prediction of solid accumulation in ASBR. In this study, results from operation A-1 were used for evaluation of the fitting results. Data before washout were used. The results were as shown in Figure 6.11.

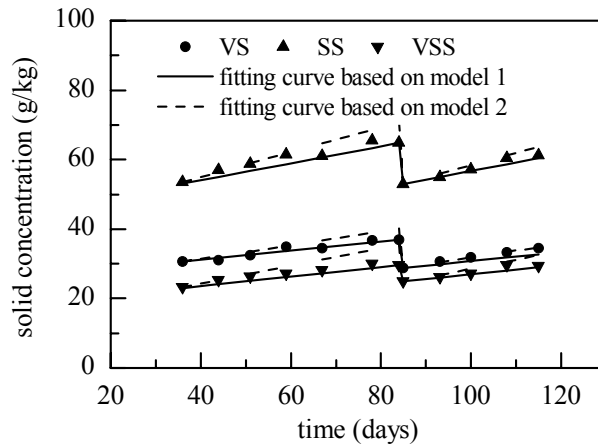


Figure 6.11 Prediction of solid accumulation in ASBR: dot- experiment data; black line- simulation data based on the model assuming hydrolysis limitation; dot line- simulation data based on the model from Romero (1991)

Although the two models showed little difference on the prediction of substrates removal rates and effluent concentration, predicted solid concentrations were higher from simulation based on Romero's model. However, both of them were acceptable with most $R^2 > 0.8$.

6.5 Conclusion

Modelling and simulation of CSTR and ASBR treating thMBW was carried out in this Chapter. SRT was used as an important parameter. A model developed from the hypothesis of hydrolysis limitation was established for prediction of substrate removal rate, effluent concentrations, and solid accumulation. Another model established by Romero (1991) was also used for simulation. Both of the simulation results were compared with the experiment data. It was found the model established in this chapter corresponded with the experiment data, and were also comparable with the simulation results from Romero's model.

It showed SRTs should be longer than about 20 days to obtain relative stable removal of the organics in MBW, which also corresponded with common suggestion from other studies. This result indicated an advantage of ASBR over CSTR, as longer SRTs can be obtained in ASBR than CSTR when they are operated under the same HRT.

However, as the amount of microorganisms and solid substrates cannot be separately determined in this study, both these two models cannot give separate description of change of these two parameters. The present results were based on indicators like 4S, which comprises both microorganisms and substrates. Further studies should separate the microorganism from substrates by other indicators.

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Chapter 7 Economic, environmental, and energetic evaluation

7.1 Introduction

Anaerobic digestion (AD) with thermal hydrolysis (TH) pre-treatment, especially application of ASBR in the digestion processes, has been proved technically acceptable for the treatment of municipal biowaste. It has been proved digestibility of the MBW can be improved through increased digestion rate. Thermal hydrolysis also improved settling performance of MBW and made application of ASBR possible. As ASBR can separate SRT from HRT, better digestion efficiency can be achieved compared with CSTR operated under the same HRTs.

Although economic, environmental, and energetic feasibility of AD technologies has been proved by many researchers (De Baere, 2000 and 2006; DiStefano and Belenky, 2009; El Hanandeh and El-Zein, 2010), its combination with TH still needs further evaluation. In the TH processes, high reaction temperature of about 150-175°C is usually suggested (Fisher and Swanwick, 1971; Haug, 1978). Correspondingly, a great deal of energy input is needed for this process, which may also increase running cost and environmental loading.

As a result, it's important to make evaluation on the energetic, economic, and environmental feasibility of this TH-AD combined technology as well. In this chapter, comparison was made among the TH-AD combined technology, single AD process, and incineration. Especially, the application of ASBR in the AD process was compared with the case using CSTR. Calculation was based on the situation in China, which may provide a reference to the situation in other areas as well.

7.2 Scenario set-up

7.2.1 Municipal biowaste

Properties and quantities of the municipal biowaste have to be determined before system evaluation. According to the experiment result, a 1/1/1 (in weight) mixture of RKW, FVW, and WSS was chosen as the target. The properties of this mixture varied during experiment.

The average results of long time monitoring (about 2 years in the pilot plant) were approximate to a TS content of 12% and VS of 10%.

In this study, digesters with an effective volume of about 0.2 m³ were used. However, the production of MBW is great and much larger treatment capacities are needed for full-scale application. As introduced in Chapter 2, Beijing already built up an AD plant with a capacity of 650 t/d, treating 350 t/d of MBW and 300 t/d of MSW. Other existing applications usually have capacities around 500-1000 t/d. According to these data, a treatment capacity (Q) of 500 t/d was chosen for the system evaluation.

Daily loading of the system can then be calculated as Table 7.1.

Table 7.1 Composition of the mixed municipal biowaste (t/d)

Q ₀	TS ₀	VS ₀	W ₀
500	60	50	440

7.2.2 Process scenarios

Figure 7.1 gives the major steps of municipal biowaste treatment. By selecting different methods for each step, various scenarios can be derived.

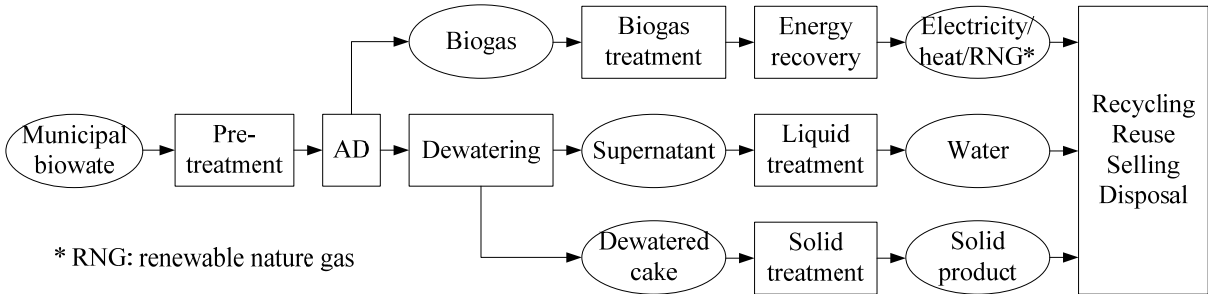


Figure 7.1 Municipal biowaste treatment process

One of the scenarios, which use thermal hydrolysis pre-treatment and ASBR digester, is shown in Figure 7.2. By changing some steps in Figure 7.2, several scenarios were developed presenting the major treatment flow discussed in this study, as shown in Table 7.2 (scenarios 1-3). In scenario 4, all the municipal biowaste were directly incinerated. This scenario was used as control, as it is one of the most commonly accepted methods for MBW treatment at present.

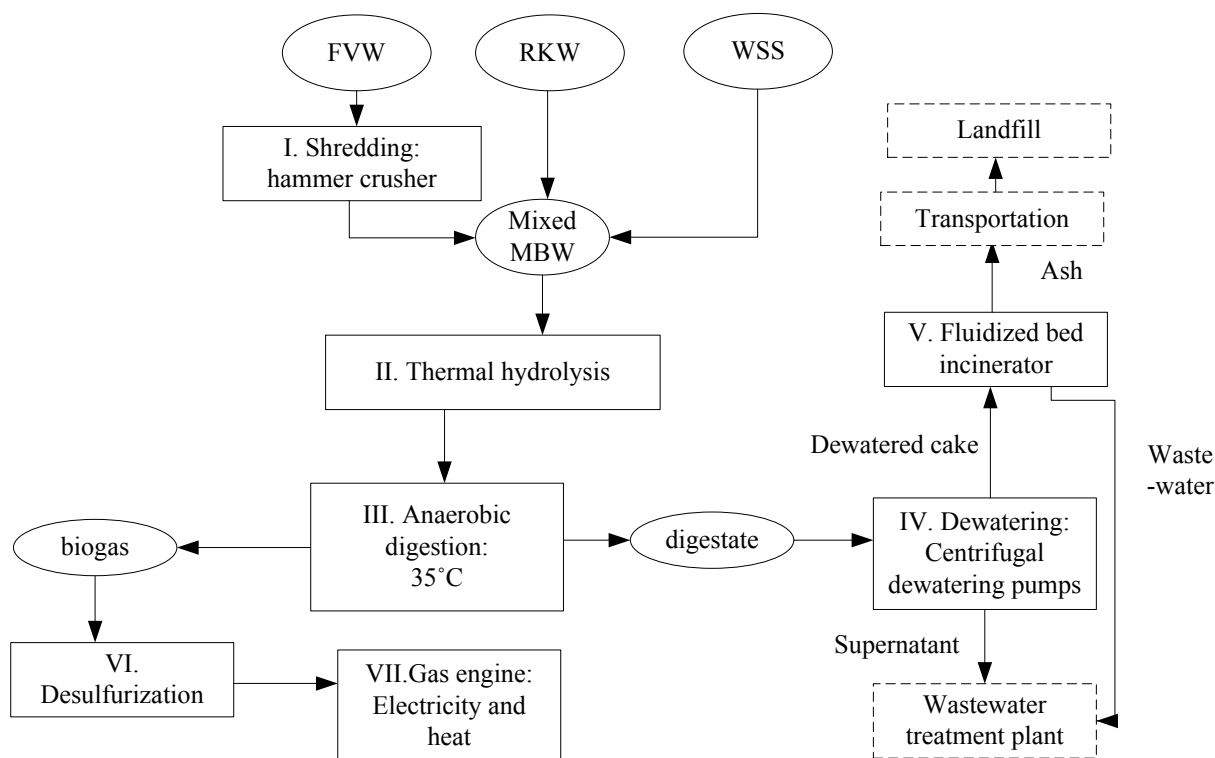


Figure 7.2 The scenario using thermal hydrolysis pre-treatment and ASBR digester

Table 7.2 Scenarios analyzed in this study

Scenarios	Processes	Description
1	I + II + III (ASBR) + IV + V + VII	TH and ASBR
2	I + II + III (CSTR) + IV + V + VII	TH and CSTR
3	I + III (CSTR) + IV + V + VI + VII	CSTR
4	V	Direct incineration (control)

In scenarios 1 and 2, desulfurization (process VI) was not taken into consideration. During the pilot experiments, H₂S was not found in the biogas when thermal hydrolysis was applied. While at the same time, average H₂S concentration of 854.9 ppm was found for biogas produced from digesters without thermal pre-treatment.

Wastewater produced in the treatment process is discharged into sewage collect system and treated by sewage treatment plants. Ash produced in the process is transported to landfill sites for landfill disposal.

7.3 Mass balance

Mass balance of each process is discussed in the following paragraphs.

7.3.1 Shredding

Shredding is needed in scenario 1, 2 and 3. Based from experience of the pilot plant, it was found only the FVW needed shredding before TH or AD treatment. As a result, the treatment capacity of shredder (Q_I) is 167 t/d. It's supposed no weight loss in this process.

7.3.2 Thermal hydrolysis

Thermal hydrolysis is used in scenario 1 and 2.

During the pilot experiments, it was found parts of the steam entered into the MBW and caused a dilution ratio of about 1.6. However, this ratio could be lower.

In the pilot plant, TH facilities were not continuously operated most of the time. As a result, the pre-heating step was not available. Wastes needed to be heated from room temperature to the reaction temperature of 175°C before TH reaction. During this heating process, large amounts of heat were needed. Parts of the steam condensed into water to release the latent heat.

If the TH facilities are operated continuously, residual heat of the treated MBW can be recovered to pre-heat the biowaste to a temperature of about 95°C. The steam is only used to heat the MBW from 95°C to 175°C. Much less steam will condense into water during this process. The dilution ratio could be reduced to about 1. In the following calculation, the dilution rate was also set as 1.

It's supposed that no TS and VS loss happened during thermal hydrolysis, which was close to experiments results shown in Chapter 3. It means the treatment capacity (Q_{II}) and waste compositions (TS_{II} , VS_{II} , W_{II}) are the same as Q_0 , TS_0 , VS_0 , and W_0 .

7.3.3 Anaerobic digestion

ASBR with TH pre-treatment, CSTR with TH, and CSTR without TH were set to be used in scenario 1, 2, and 3.

From the results in Chapter 6, it was found ASBR showed significant advantages over CSTR when HRTs were shorter than about 20 d. However, its advantage may not be so

obvious when longer HRTs were selected. Accordingly, results from operation A-2 and C-2 with a HRT of 18 d, were used as references in scenario 1 and 2. CSTR without TH were not operated under this HRT in this study. However, results from a former study in the Prof.Wang's group (Liu, 2012) were used. Their performances were listed in Table 7.3.

Table 7.3 Operating conditions and performance of the digesters

Indicators	Description	Scenario 1	Scenario 2	Scenario 3
HRT	HRT (d)	18	18	18
Y	Biogas yield (Nm ³ /t-VS _{added})	801	680	608
C _M	Methane content (% in volume)	69.3	68.8	55.2
C _C	CO ₂ content (% in volume)	30.7	31.2	44.0
C _S	H ₂ S content (% in volume)	0	0	0.8

It's assumed that all the organics removed turn into biogas. No inorganic matters loss happens. Based on it, mass balance can be made as Table 7.4.

Table 7.4 Mass balance in the AD process

input	output [#]	
	biogas	digestate
$Q_{III}=Q_0$	$Q_{\text{biogas}}=\sum VS_{III} * Y * C_i * M_i / 22.4 / 1000$	$Q_{\text{digestate}}=Q_{III}-\sum VS_{III} * Y * C_i * M_i / 22.4 / 1000$
$TS_{III}=TS_0$	$TS_{\text{biogas}}= Q_{\text{biogas}}$	$TS_{\text{digestate}}=TS_{III} - TS_{\text{biogas}}$
$VS_{III}=VS_0$	$VS_{\text{biogas}}= Q_{\text{biogas}}$	$VS_{\text{digestate}}=VS_{III} - VS_{\text{biogas}}$
$W_{III}=W_0$	0	$W_{\text{digestate}}=W_{III}$

[#]_i means different contents of the biogas, which are CH₄, CO₂, H₂S in this study; M_i means the molar mass of each component, g/mol.

7.3.4 Dewatering

In chapter 4, it was found dewatering performance of the waste could be improved after TH. Based on the results listed in Table 4.7, water content of the dewatered digestate cake was reduced from about 80% to 60% when TH was applied. VS contents of the digestate were 53.2 ± 0.8%, 58.2 ± 1.4%, and 63.2% for according operations used in scenario 1, 2, and 3.

Accordingly, performance of the dewatering process in scenarios 1-3 is set close to the above experimental data, as shown in Table 7.5. Addition of flocculants and/or other chemicals is not considered under all these scenarios.

Table 7.5 Dewatering performance

indicator	description	scenario 1	scenario 2	scenario 3
R_{TS}	TS recovery rate (%)	85	85	85
f_w	Water content of dewatered cake (%)	60	60	80
f_{vs}	VS/TS (%)	50	60	60

Mass distribution in dewatered cake and the separated water can then be calculated from these data. Results are listed in Table 7.6.

Table 7.6 Mass balance in the dewatering process

input	output	
	dewatered cake	separated water
$Q_{IV}=Q_{digestate}$	$Q_c=TS_{IV} * R_{TS}/(1-f_w)$	$Q_{sw}=Q_{IV}-Q_c$
$TS_{IV}=TS_{digestate}$	$TS_c=TS_{IV} * R_{TS}$	$TS_{sw}=TS_{IV}-TS_c$
$VS_{IV}=VS_{digestate}$	$VS_c=TS_{IV} * f_{vs}$	$VS_{sw}=VS_{IV}-VS_c$

7.3.5 Incineration

Fluidized-bed incinerator is supposed to be used in the incineration process. The treatment flow is shown as Figure 7.3, which follows the suggestion of Murakami (1988).

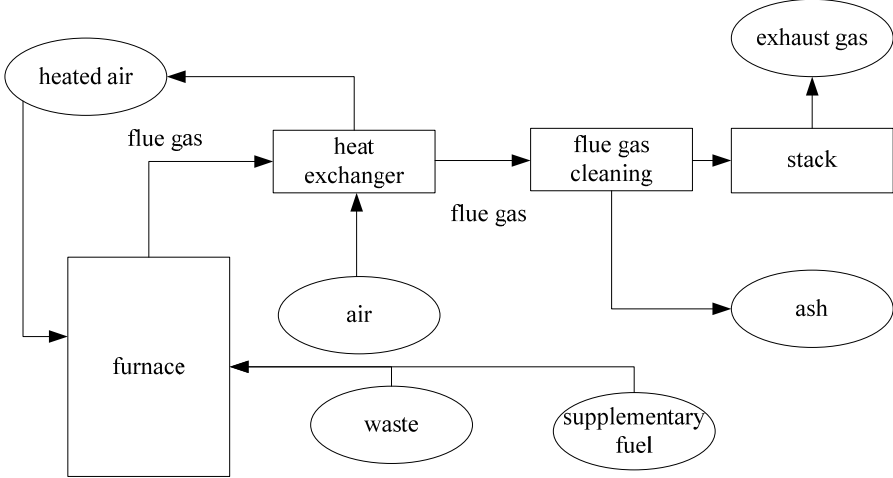


Figure 7.3 Treatment steps of the incineration process

The waste treated is the dewatered cake of digestate under scenario 1-3 and the MBW under scenario 4. Amounts of the digestate can be calculated from Table 7.6. Table 7.7 gives constituents of the waste under each scenario. Most data were from experiment results of the pilot plant. However, element analysis was not carried out for scenario 2 and it was considered as the same as scenario 1.

Table 7.7 Constituents of the waste

Indicators	Scenario 1	Scenario 2	Scenario 3	Scenario 4
C (%-VS)	50.38	50.38	45.93	48.43
H(%-VS)	6.80	6.80	6.96	7.43
N(%-VS)	5.84	5.84	7.47	5.66
O(%-VS)	35.21	35.21	37.94	37.55
S(%-VS)	1.77	1.77	1.70	0.92

Constituents of air are listed in Table 7.8. In scenario 3 and 4, fuel supplement is found needed. In scenario 3, biogas produced from the process of AD is used as supplementary fuel. Its constituents have been listed in Table 7.3. In scenario 4, coal is used, which is relatively cheap and commonly used in China. Its constituents are listed in Table 7.8.

Table 7.8 Constituents of air and coal

constituent	air (mole fraction)	constituent	coal (weight fraction)
CO ₂	0.0003	C	0.612
N ₂	0.7802	H	0.043
O ₂	0.2069	N	0.012
SO ₂	0	O	0.074
moisture	0.0126	S	0.039
density (kg/Nm ³)	1.28	ash (A _c)	0.120
		moisture (W)	0.100

Major reactions during the incineration process are as following:





Based on the constituents' results and the above equation, air required for combustion can be calculated out. 30% of excess combustion air is considered (Murakami, 1988).

$$D_{\text{air}} = 1.3 \times (C/12 + H/4 + S/32.1 - O/16/2) / 0.2069 \quad (7-6)$$

where C, H, O means the weight fraction of carbon, hydrogen, and oxygen in the combustible components of waste or fuel; D_{air} means the demand of air, mol/g-waste or fuel. All these indicators are marked with subscript w or f for the expression of properties of waste or fuel.

Quantity of combustion air (Q_{air}) can be calculated out if the amounts of waste and fuel are known.

$$Q_{\text{air},w} (\text{air required for waste combustion, t/d}) = D_{\text{air}} \cdot M_{\text{air}} \cdot Q_w \quad (7-7)$$

$$Q_{\text{air},f} (\text{air required for fuel combustion, t/d}) = D_{\text{air}} \cdot M_{\text{air}} \cdot F \quad (7-8)$$

where M_{air} means the molecular weight of air, which is 28.706 g/mol; Q_w means the quantity of waste, t/d; F means the quantity of supplementary fuel, t/d.

Quantity of the flue gas (Q_f) can then be calculated out following Table 7.9. It's also marked with subscript w or f for the expression of flue gas produced from waste or fuel.

Table 7.9 Flue gas production and composition

constituent	flue gas (mol/g-waste or fuel)	references
CO_2	$C/12 + 0.0003 \cdot D_{\text{air}}$	
N_2	$N/14/2 + 0.7802 \cdot D_{\text{air}}$	
O_2	$D_{\text{air}}/1.3 \cdot 0.3 \cdot 0.2069$	
SO_2	$S/32.1$	
moisture (W_f)	$H/2 + W/18 + 0.0126 \cdot D_{\text{air}}$	
ash (A, t/d)	waste: $(\text{TS}-\text{VS}) \cdot Q_w$; fuel: $A_c \cdot F$	TS, VS: contents in waste, %

Quantity of supplementary fuel (F) can be calculated from heat balance.

Table 7.10 gives some indicators used in the heat balance process.

Table 7.10 Indicators used for heat balance in the incineration process

indicator	description	value	unit	reference
HV _w	gross heating value of the wastes	by calculation		Boie (1952)
HV _m	gross heating value of methane	39.8	MJ/Nm ³	
HV _c	gross heating value of coal	25.92	MJ/kg	
C _{p,a}	specific heat of air	by calculation	kJ/kg/°C	interpolation
C _{p,f}	specific heat of flue gas (without ash)	by calculation	kJ/kg/°C	interpolation
C _{p,ash}	specific heat of ash	0.84	kJ/kg/°C	
C _l	latent heat of water	2257	kJ/kg	
T ₀	ambient temp. (set as datum)	20	°C	Murakami (1988)
T ₁	flue gas temp. at the outlet of furnace	850	°C	Murakami (1988)
T ₂	air temp. at the outlet of heat exchanger	650	°C	Murakami (1988)
T ₃	flue gas temp. at the outlet of heat exchanger	by calculation		heat balance
η ₁ [#]	heat loss ratio in furnace	29.6066Q ^{-0.43697}	%	Murakami (1988)
η ₂	heat loss ratio in heat exchanger	10	%	Murakami (1988)

#Q means treatment capacity of the furnace, t/d.

Among them, gross heating value of the wastes (HV_w) is calculated following Boie's formula (1952), which is as:

$$HV_w = 35160 C' + 116225 H' - 11090 O' + 6280 N' + 10465 S' \quad (7-9)$$

where HV_w means the gross heating value of the wastes, kJ/kg-TS; C', H', O', N', S' means the weight fraction of these elements in dry basis.

Specific heat values of CO₂, SO₂, H₂O, N₂, and O₂ under a certain temperature are calculated by interpolation. Specific heat of air (C_{p,a}) and flue gas (C_{p,f}) is then calculated based on their contents of these constituents.

Heat balance is then carried out using these indicators. Heat input to the furnace can be calculated as Table 7.11, while heat output from the furnace can be calculated as Table 7.12.

Table 7.11 Heat input to furnace

indicators	description	heat input
Q01	combustion heat of waste	$HV_w * Q_w$
Q02	combustion heat of fuel	$HV_f * F$
Q03	sensible heat of air used for waste combustion	$Q_{air,w} * (T_2 - T_0) * C_{p,a}$
Q04	latent heat of air used for waste combustion	$Q_{air,w} / M_{air} * 0.0126 * 18 * C_1$
Q05	sensible heat of air used for fuel combustion	$Q_{air,f} * (T_2 - T_0) * C_{p,a}$
Q06	latent heat of air used for fuel combustion	$Q_{air,f} / M_{air} * 0.0126 * 18 * C_1$
Q0	heat input to furnace	$\sum Q0i$

Table 7.12 Heat output from furnace

indicators	description	heat output
Q11	sensible heat of flue gas from waste combustion (without ash)	$Q_{f,w} * (T_1 - T_0) * C_{p,f}$
Q12	latent heat of flue gas from waste combustion (without ash)	$W_{f,w} * Q_w * 18 * C_1$
Q13	sensible heat of flue gas from fuel combustion (without ash)	$Q_{f,f} * (T_1 - T_0) * C_{p,f}$
Q14	latent heat of flue gas from fuel combustion (without ash)	$W_{f,f} * Q_w * 18 * C_1$
Q15	heat of ash produced from waste combustion	$A_w * (T_1 - T_0) * C_{p,ash}$
Q16	heat of ash produced from fuel combustion	$A_f * (T_1 - T_0) * C_{p,ash}$
Q17	heat loss	$Q0 * \eta_1$
Q18	excess heat	$(Q11 + Q12 + \dots + Q17) - Q0$
Q1	heat output from furnace	$\sum Q1i$

When $Q18 < 0$, fuel supplement is needed. F can then be obtained by tentative calculation, which making $Q18 = 0$.

Based on it, heat balance in the heat exchanger can also be made. Heat input to heat exchanger (Q2) can be calculated as:

$$Q2 = Q11 + Q12 + Q13 + Q14 + Q15 + Q16 \quad (7-10)$$

Heat output from the heat exchanger can be calculated as Table 7.13. Among them, T_3 is obtained by repeating tentative calculation until a balance between Q2 and Q3 is made.

Table 7.13 Heat output from heat exchanger

indicators	description	heat input
Q31	sensible heat of flue gas from waste combustion(without ash)	$Q_{f,w}*(T_3-T_0)*C_{p,f}$
Q32	latent heat of flue gas from waste combustion (without ash)	$W_{f,w}*Q_w*18*C_l$
Q33	sensible heat of flue gas from fuel combustion(without ash)	$Q_{f,f}*(T_3-T_0)*C_{p,f}$
Q34	latent heat of flue gas from fuel combustion (without ash)	$W_{f,f}*Q_w*18*C_l$
Q35	heat of ash produced from waste combustion	$A_w*(T_3-T_0)*C_{p,ash}$
Q36	heat of ash produced from fuel combustion	$A_f*(T_3-T_0)*C_{p,ash}$
Q37	heat used for exchange	$Q03+Q04+Q05+Q06$
Q38	heat loss	$Q37*\eta_2$
Q3	heat output from heat exchanger	$\sum Q3i$

The waste heat like excess heat (Q18) and heat after exchanger (Q3-Q37-Q38), is supposed be utilized by a waste heat boiler. This part of heat can be used for heating in the process of TH and AD. Heat loss of the boiler is set as 6% (Murakami, 1988).

7.3.6 Desulfurization

Biogas desulfurization is needed only in scenario 3. Biogas quantity and its H_2S content can be calculated from Table 7.3 and 7.4. It's supposed all the H_2S is removed in this process.

Under scenario 3, a part of the biogas is used as supplementary fuel in the incineration process. Its amount can be calculated from heat balance in the section of 7.3.5. Mass distribution in this process can then be calculated as Table 7.14.

Table 7.14 Mass balance in the desulfurization process (scenario 3 only)

input	output
$Q_{VI}=Q_{biogas}-F$	$Q_{VII}=Q_{VI}-VS_0*Y*Cs*M_{H_2S}/22.4/1000$
including: $CO_2: Q_{VI}*C_{CO_2}$	including: $CO_2: Q_{VI}*C_{CO_2}$
$CH_4: Q_{VI}*C_{CH_4}$	$CH_4: Q_{VI}*C_{CH_4}$
$H_2S: Q_{VI}*C_{H_2S}$	$H_2S: 0$

7.3.7 Biogas utilization by gas engine

Gas engine is used to recover energy from biogas. In scenario 1 and 2, all the biogas produced is treated by biogas engine. Calculation of their amounts has been introduced in Table 7.4 ($Q_{VII} = Q_{III}$). In scenario 3, amount of biogas treated by gas engine is as shown in Table 7.14 (Q_{VII}).

It's supposed all the methane is burned into CO_2 and H_2O in this process. Air is supplied for oxygen demand. Its properties have been listed in Table 7.8. The quantity of air supply and exhaust gas is as following:

Table 7.15 Mass balance for gas engine

	input		output
	biogas	air	
CH ₄	$Q_{CH_4} = Q_{VII} \cdot C_{CH_4}$ or $Q_{VI} \cdot C_{CH_4}$ (scenario 3)	0	0
CO ₂	$Q_{CO_2} = Q_{VII} \cdot C_{CO_2}$ or $Q_{VI} \cdot C_{CO_2}$ (scenario 3)	$Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.0003 \cdot 44$	$Q_{CH_4}/16 \cdot 44 + Q_{CO_2} +$ $Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.0003 \cdot 44$
H ₂ O	0	$Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.0126 \cdot 18$	$Q_{CH_4}/16 \cdot 2 \cdot 18 +$ $Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.0126 \cdot 18$
N ₂	0	$Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.7802 \cdot 28$	$Q_{CH_4}/16 \cdot 2/0.2069 \cdot 0.7802 \cdot 28$
O ₂	0	$Q_{CH_4}/16 \cdot 2 \cdot 32$	0

7.4 Energy consumption

7.4.1 Shredding

In the experiment, they were shredded by a hammer crusher (A300, BestPower, China) to an average size of less than 3.0 mm. Its power efficiency (f_{PI}) is 7.3 kWh/t. As electricity is a source of secondary energy, the efficiency of electricity generation (f_E) should be considered. It's set as 0.355 (Shomura, 2010).

Energy consumption (E_I , MJ/d) can be calculated as:

$$E_I = 3.6 \times f_{PI} \cdot Q_I / f_E \quad (7-11)$$

7.4.2 Thermal hydrolysis

Energy input is needed both for heating and power consumption of the pumps. Power efficiency of pumps (f_{pII}) was obtained from experimental data, which is 0.289 kWh/t.

Heat need (H_{II} , MJ/d) can be calculated as:

$$H_{II} = C_p \cdot Q_{II} \cdot (T_{II,r} - T_{II,i}) \quad (7-12)$$

Description and values of indicators shown in the above equation are listed in Table 7.16.

Table 7.16 Indicators for the process of TH

indicators	description	value	unit	references
Q_{II}	treatment capacity	500	t/d	set value
$T_{II,i}$	initial temperature	95	°C	experimental data
$T_{II,r}$	reaction temperature	175	°C	experimental data
C_p	specific heat of MBW at a constant pressure	4.1868	kJ/(kg·°C)	Shomura, 2010
f_b	efficiency of electric boiler	0.95	-	set value

Heat is supposed to be provided by waste heat generated from incineration and heat recovered from gas engine. If these two sources of heat can't cover the heat demand, electric boiler is supposed to be used. Electricity demand (P_{II} , MJ/d) can be calculated under this condition as:

$$P_{II} = (H_{II} - (-H_{VII} + Q_{18} + Q_{37} - Q_{38}) \cdot (1-6\%)) / f_b \quad (7-13)$$

where $-H_{VII}$ means heat produced by gas engine, which will be introduced in section 7.4.7.

The total energy consumption (E_{II} , MJ/d) can then be calculated as:

$$E_{II} = (3.6 \times f_{pII} \cdot Q_{II} + P_{II}) / f_E + (H_{II} - P_{II} \cdot f_b) \quad (7-14)$$

7.4.3 Anaerobic digestion

For scenario 1 and 2, thermally hydrolyzed MBW has a high residual temperature. As a result, heat input isn't needed in AD process.

In scenario 3, heat input (H_{III}) can be calculated as following:

$$H_{III} = C_p \cdot Q_{III} \cdot (T_{III,r} - T_{III,i}) / f_h \quad (7-15)$$

The related indicators are listed in Table 7.17.

Table 7.17 Indicators for heating in the AD process

indicators	description	value	unit	references
Q_{III}	treatment capacity	500	t/d	set value
$T_{III,i}$	initial temperature	20	°C	set value
$T_{III,r}$	reaction temperature	35	°C	set value
f_h	heating efficiency	0.7	-	Shomura, 2010

Heat is supposed to be firstly provided by heat produced by gas engine ($-H_{VII}$). If $H_{III} > -H_{VII}$, coal is used as supplement. Its consumption (F_{III} , kg/d) can be calculated as:

$$F_{III} = (H_{III} - H_{VII}) / HV_c \quad (7-16)$$

Extra energy is needed for pumping, mixing, and so on in scenario 1-3. Its amount is calculated based on an equation developed by Shomura (2010).

$$f_{pIII} = 2587.2 x_{III}^{-0.9439} \quad (7-17)$$

where f_{pIII} means the electricity efficiency in the AD facilities, kWh/t-TS; x_{III} means the treatment capacity of the AD facilities, t-TS/d.

Energy consumption (E_{III} , MJ/d) can then be calculated as:

$$E_{III} = 3.6 \times f_{pIII} \cdot x_{III} / f_E + H_{III} \quad (7-18)$$

7.4.4 Dewatering

Centrifugal drier is selected for dewatering. Its electricity efficiency (f_{pIV} , kWh/t-TS) can be calculated from an equation developed by Shomura (2010).

$$f_{pIV} = 885.53 x_{IV}^{-0.2989} \quad (7-19)$$

where x_{IV} means the treatment capacity of the drier, t-TS/d.

Energy consumption (E_{IV} , MJ/d) can then be calculated as:

$$E_{IV} = 3.6 \times f_{pIV} \cdot x_{IV} / f_E \quad (7-20)$$

7.4.5 Incineration

Electricity efficiency (f_{pV} , kWh/t-TS) in the incineration process can be calculated from an equation developed by Shomura (2010).

$$f_{pV} = 1074.8x_V^{-0.3232} \quad (7-21)$$

where P_{IV} means the electricity consumption in the incineration process, kWh/t-TS; x_V means the treatment capacity of incineration, t-TS/d.

Coal is used in scenario 4 as supplementary fuel. Energy brought by it is expressed as Q02 in Table 7.11.

Energy consumption (E_V , MJ/d) can then be calculated as:

$$E_V = 3.6 \times f_{pV} \cdot TS_V / f_E \quad (\text{scenario 1-3}) \quad (7-22)$$

$$E_V = 3.6 \times f_{pV} \cdot TS_V / f_E + Q02 \quad (\text{scenario 4}) \quad (7-23)$$

where TS_V equals to TS_c as listed in Table 7.6 under scenario 1-3; for scenario 4, it equals to TS_0 as listed in Table 7.1.

7.4.6 Desulfurization

In scenario 3, desulfurization is needed. However, a part of the biogas produced from the AD process is used as supplementary fuel in the incineration process. This part of biogas doesn't need desulfurization. The volume of biogas treated can be calculated as:

$$V_{VI} = VS_0 \cdot Y - F_{\text{in volume}} \quad (7-24)$$

where $F_{\text{in volume}}$ can be calculated from F , which can be obtained from heat and mass balance in the incineration process, Nm^3/d .

Electricity efficiency for biogas desulfurization (f_{pVI}) is set as 0.223 kWh/ Nm^3 biogas according to experimental data obtained in Prof. Takaoka's group. Energy consumption (E_{VI} , MJ/d) can then be calculated as:

$$E_{VI} = 3.6 \times f_{pVI} \cdot V_{VI} / f_E \quad (7-25)$$

where TS_V equals to TS_c as listed in Table 7.6 under scenario 1-3; for scenario 4, it equals to TS_0 as listed in Table 7.1.

7.4.7 Biogas utilization by gas engine

It's supposed 30% of the energy in biogas can be recovered in the form of electricity and 40% of the energy in biogas can be recovered in the form of heat. Recovered heat is used for heating in the AD process under scenario 3.

$$E_{VII} = -0.7 \times VS_0 \cdot Y \cdot C_M / 100 \cdot HV_m \quad (7-26)$$

$$H_{VII} = -0.4 \times VS_0 \cdot Y \cdot C_M / 100 \cdot HV_m \quad (7-27)$$

7.5 Running cost

Running cost includes cost for electricity, fuels, chemicals, and also final disposal.

7.5.1 Electricity fee

Electricity generated from gas engine is used to cover the electricity demand in the system. If it's not enough, extra power input is needed. If excess electricity exists, it's supposed to be sold to the electricity grid.

Electricity prices are based on the situation in Beijing. According to a notice from the National Development and Reform Commission (NDRC) of China (2013), on-grid electricity price produced from a MSW treatment plant was set as 0.595 CNY/kWh. This data was used as on-grid electricity price in this chapter as well. Meanwhile, the retail electricity price for industrial purpose (<1kV) is 0.781 CNY/kWh. With an exchange rate of 627.96 /100 (Bank of China, 2013), electricity prices in US dollar was calculated in Table 7.18.

Table 7.18 Electricity prices

indicators	description	value	references
Pe,r	retail electricity price (\$/kWh)	0.124	NDRC, 2013
Pe,o	on-grid electricity price (\$/kWh)	0.095	NDRC, 2013

7.5.2 Cost for fuels

Coal and gasoline is used as fuel. Price of coal is set as the international price (Infomine, 2013), which is 62.28 \$/t-standard coal. Standard coal means coals with gross heating value of 31.52 MJ/kg. Gasoline in Beijing is 1.323 \$/l (Prices of refined oil Network, 2012).

Coal is used for heating in the AD process under scenario 3 and in the incineration process under scenario 4. Its demand during incineration can be calculated from mass and heat balance as discussed before.

Gasoline is selected as fuel used for transportation of ash. Its consumption (P, l/d) can be calculated as:

$$P = f_{\text{gasoline}} \cdot L \cdot (1 + W_{\text{ash}}) \cdot (A_w + A_f) / (D_{\text{ash}} \cdot T) \quad (7-28)$$

Related indicators are described as Table 7.19.

Table 7.19 Indicators related to transportation (Shomura, 2010)

indicator	description	value
W_{ash}	water added to ash before transportation (kg/kg-ash)	0.2
D_{ash}	stacking density of ash (t/m ³)	0.6
L	transportation distance (km)	100
T	truck capacity (m ³)	20
f_{gasoline}	fuel efficiency (gasoline, l/km)	2.5

7.5.3 Cost for chemical dosage

Chemical dosage is needed for desulfurization and incineration. 25% NaOH solution is used for biogas desulfurization and flue gas cleaning. Quartz sand and water are needed in incineration. Their price and consumption are listed in Table 7.20. Prices of quartz sand and NaOH solution were from references. The exchange rate of JPY/USD is set as 81/1 .

Table 7.20 Indicators related to chemical dosage

	indicator	value	references
unit price	quartz sand (\$/kg)	0.37	Shomura, 2010
	water(\$/m ³)	0.99	Beijing Water Bureau, 2012
	25% NaOH solution(\$/kg)	0.17	Shomura, 2010
consumption	water for incineration (m ³ /t)	38	Shomura, 2010
	25% NaOH for incineration (kg/t-TS)	57.5	Shomura, 2010
	quartz sand for incineration (kg/t)	6	Shomura, 2010
	25% NaOH for desulfurization (t/t-H ₂ S)	0.107	calculated by reaction formula

7.5.4 Cost for final disposal

Final disposal is needed for both ash and wastewater. Ash is supposed to be treated by landfill, while the separated water was treated by wastewater treatment plant. The cost of landfill is based on studies in Japan, which is set as 98.8\$/t (Shomura, 2010). Wastewater is produced from the processes of incineration and dewatering, including the separated water and water used for incinerator. Cost for wastewater treatment is based on the sewage discharging fee in Beijing, which is 0.17\$/m³ (Beijing Water Bureau, 2012).

7.6 Greenhouse gas emission

Greenhouse gas (GHG) emission is considered for the environmental evaluation of the system. CO₂ emission factors are listed in Table 7.21. CH₄ and NO_x mission factors in each treatment process are listed in Table 7.22. CH₄ and NO_x mission factors in processes not listed in Table 7.22 are set as 0.

Table 7.21 CO₂ emission factors

subjects	value	unit	references
electricity	0.839	kg-CO ₂ /kWh	EIA, 2007
coal	2.41	kg-CO ₂ /kg	IPCC, 2006
gasoline	2.62	kg-CO ₂ /L	IPCC, 2006
25% NaOH solution	0.339	kg-CO ₂ /kg	IPCC, 2006
industrial water	0.11	kg-CO ₂ /m ³	Shomura, 2010

Table 7.22 CH₄ and N₂O emission factors

Process	CH ₄		N ₂ O		references
shredding		0.00908	kg-CO ₂ eq./kg		CFP, 2011
dewatering	48	g-CH ₄ /t-TS	0.2	g-N ₂ O/t-TS	IPCC, 2006
incineration	0	g-CH ₄ /t	67	g-N ₂ O/t	IPCC, 2006
GWP	21	kg-CO ₂ eq./kg-CH ₄	310		IPCC, 2006

* SASP: standard activated sludge processes

7.7 Results and discussion

7.7.1 Mass balance

Results of mass balance are listed in Figure 7.4-7.7.

During mass balance, it's important to make a heat balance in the process of incineration. The heat balance helps to calculate the amount of supplementary fuels needed, and also other parameters like the amounts of combustion air, flue gas, and so on. Heat balance results under each scenario are shown in Table 7.23-7.26.

It's calculated that T_3 is 510.4°C, 420.6°C, 536.3°C, and 536.1°C for scenario 1, 2, 3, and 4.

7.7.2 Energy consumption

Based on mass balance results, energy consumption of each process can be calculated out. Results are shown in Figure 7.8.

It can be found, energy consumption is reduced in scenario 1-3 compared with scenario 4. Most of the reduction results from reduced treatment capacity of the incineration process. As parts of the waste turn into biogas and most of the water it contains is separated by dewatering, incineration capacity is greatly decreased in scenario 1-3.

Consequently, the whole system became energy-producing in scenario 1-3. The energy produced is: scenario 1>2>3. It mainly results from improvement of biogas production by TH and ASBR, and also the dewaterability improved by this process/technology.

7.7.3 Running cost

Running cost includes cost for electricity, fuels, chemicals, and also final disposal. The results are as shown in Figure 7.9.

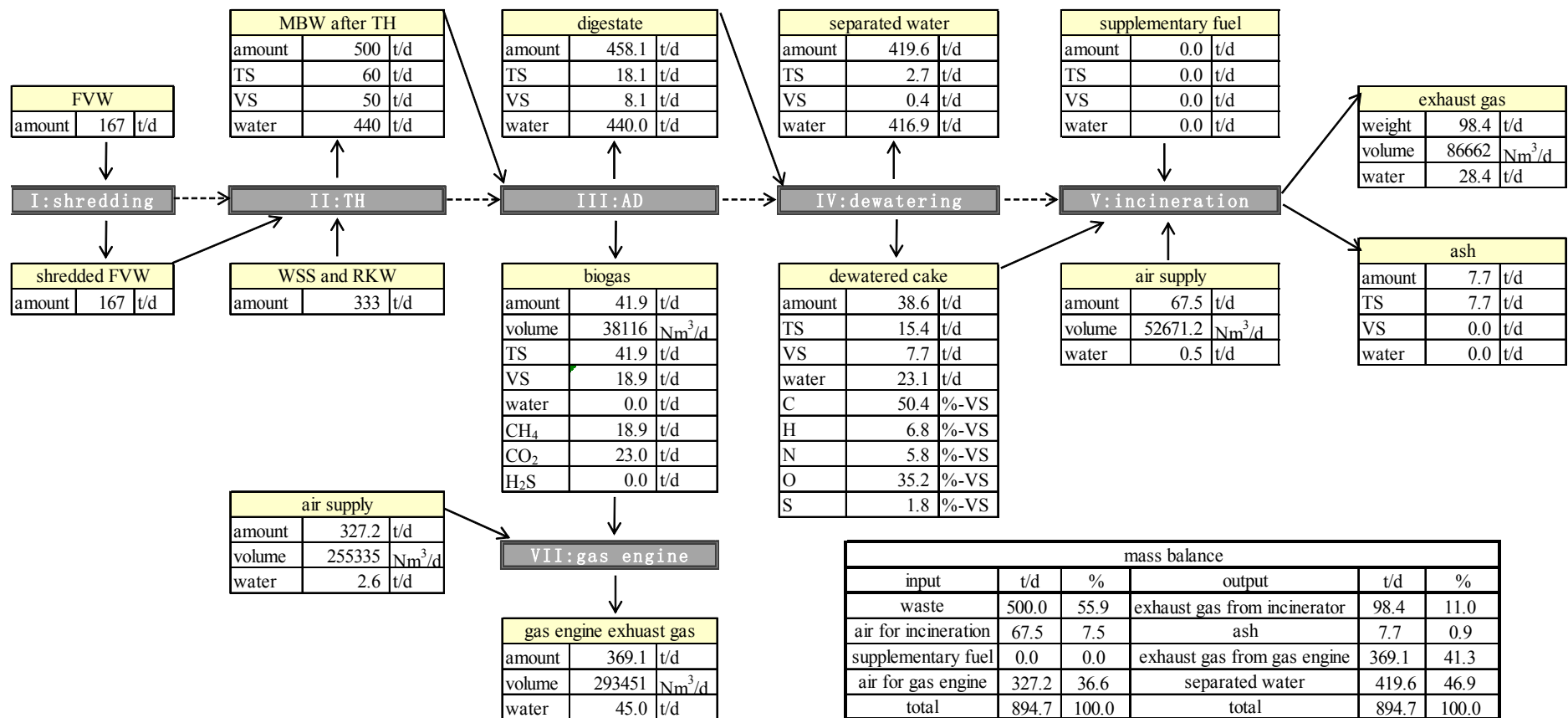


Figure 7.4 Mass balance in scenario 1

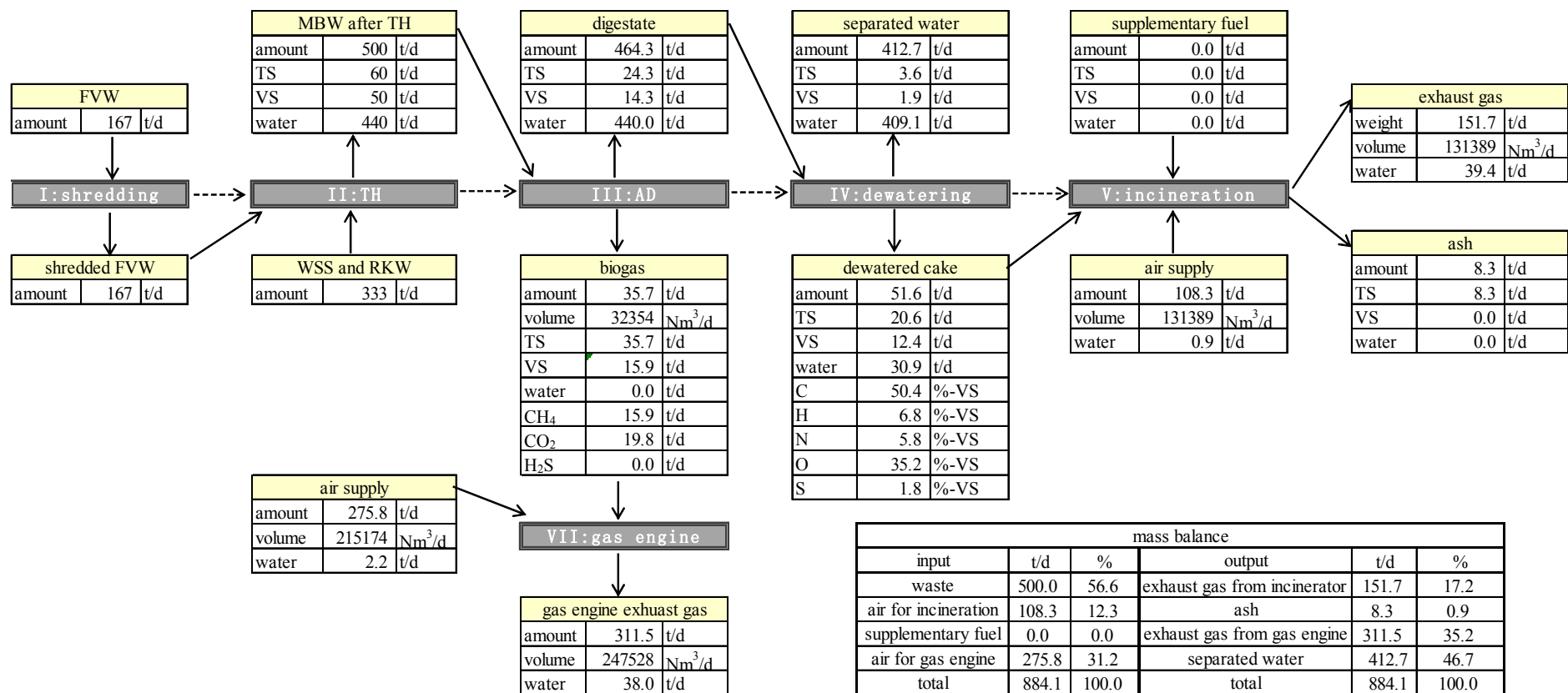


Figure 7.5 Mass balance in scenario 2

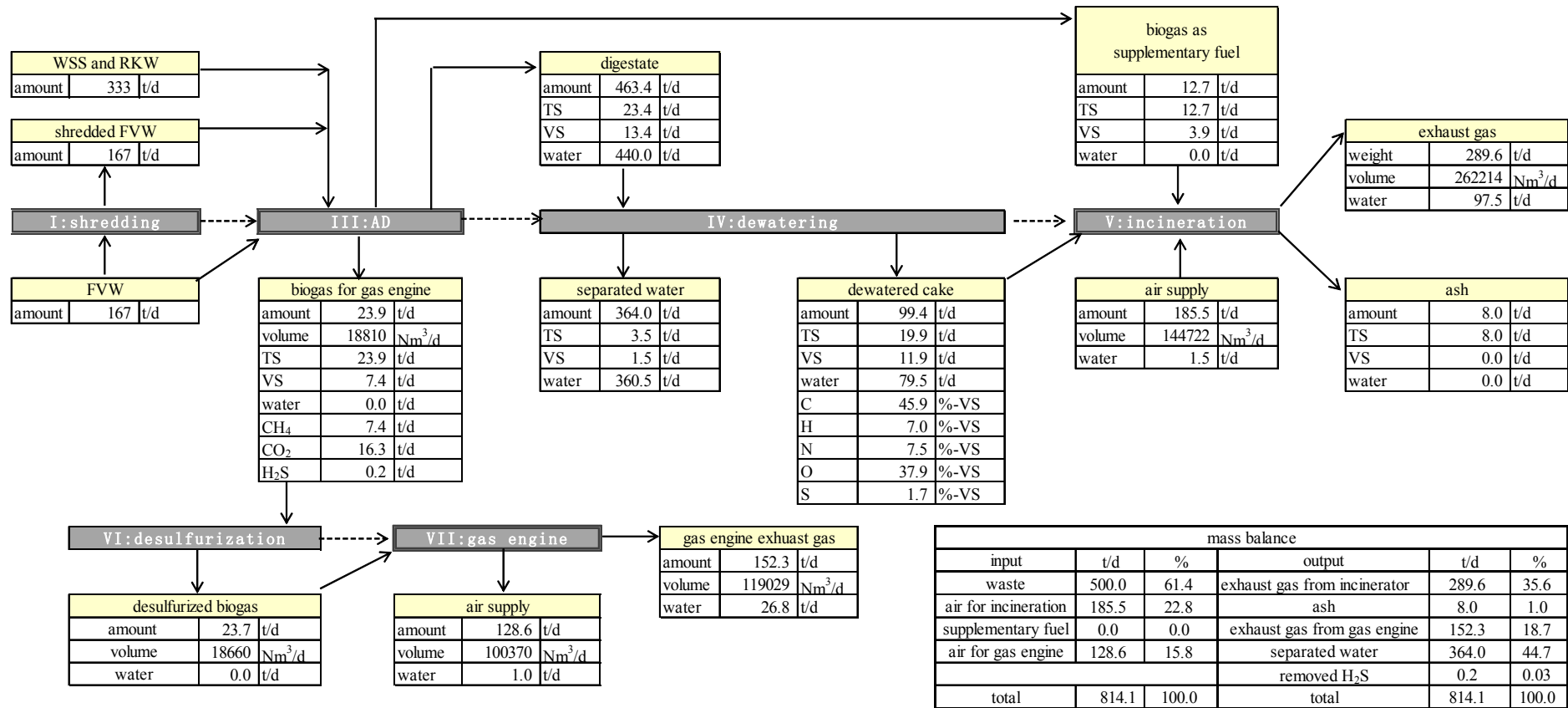
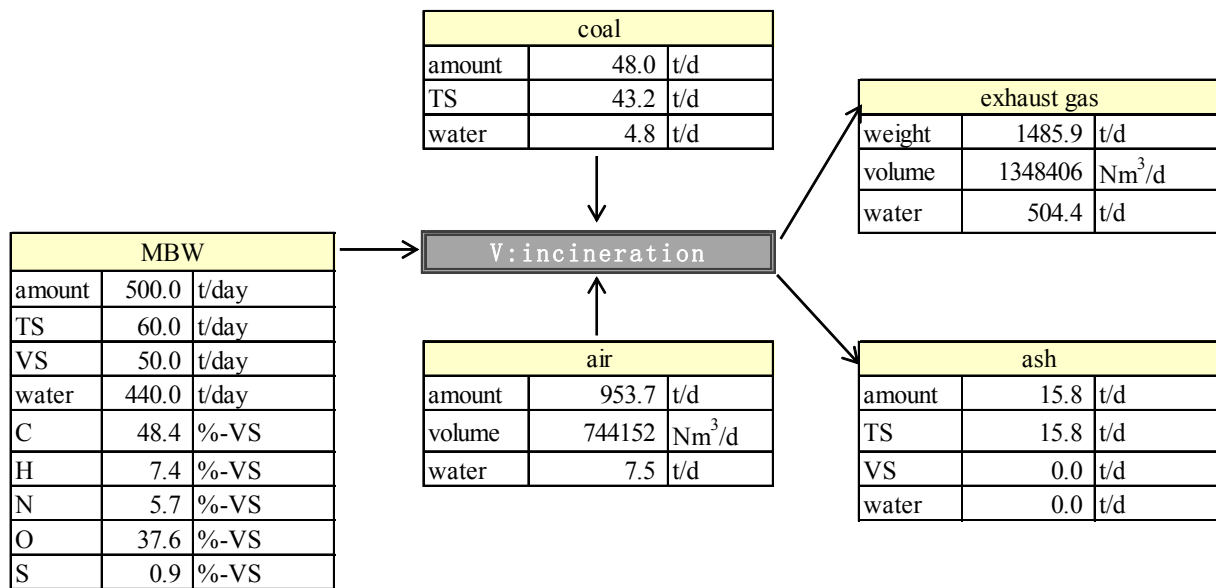


Figure 7.6 Mass balance in scenario 3



mass balance					
input	t/d	%	output	t/d	%
waste	500.0	33.3	exhaust gas	1485.9	99.0
air	953.7	63.5	ash	15.8	1.0
supplementary fuel	48.0	3.2			
total	1501.7	100.0	total	1501.7	100.0

Figure 7.7 Mass balance in scenario 4

Table 7.23 Heat balance in the incineration process under scenario 1

	MJ/d	%		MJ/d	%		MJ/d	%
Q01	171681	78.7	Q11	113265	51.9	Q31	64210	35.1
Q02	0	0.0	Q12	64075	29.4	Q32	64075	35.1
Q03	45392	20.8	Q13	0	0.0	Q33	0	0.0
Q04	1204	0.6	Q14	0	0.0	Q34	0	0.0
Q05	0	0.0	Q15	5377	2.5	Q35	3177	1.7
Q06	0	0.0	Q16	0	0.0	Q36	0	0.0
			Q17	13307	6.1	Q37	46595	25.5
			Q18	22253	10.2	Q38	4660	2.6
Q0	218277	100	Q1	218277	100			
heat exchanger			Q2	182717	100	Q3	182717	100

Table 7.24 Heat balance in the incineration process under scenario 2

	MJ/d	%		MJ/d	%		MJ/d	%
Q01	275545	78.7	Q11	170925	48.8	Q31	91639	34.5
Q02	0	0.0	Q12	88870	25.4	Q32	88870	33.5
Q03	72853	20.8	Q13	0	0.0	Q33	0	0.0
Q04	1932	0.6	Q14	0	0.0	Q34	0	0.0
Q05	0	0.0	Q15	5754	1.6	Q35	2777	1.0
Q06	0	0.0	Q16	0	0.0	Q36	0	0.0
			Q17	18832	5.4	Q37	74785	28.2
			Q18	65949	18.8	Q38	7478	2.8
Q0	350330	100	Q1	350330	100			
heat exchanger			Q2	265548	100	Q3	265548	100

Table 7.25 Heat balance in the incineration process under scenario 3

	MJ/d	%		MJ/d	%		MJ/d	%
Q01	247193	41.6	Q11	244134	41.0	Q31	145708	25.5
Q02	219661	36.9	Q12	198178	33.3	Q32	198178	34.7
Q03	64301	10.8	Q13	101185.17	17.0	Q33	60876	10.7
Q04	1705	0.3	Q14	21766.87	3.7	Q34	21767	3.8
Q05	60420	10.2	Q15	5545	0.9	Q35	3450	0.6
Q06	1602	0.3	Q16	0	0.0	Q36	0	0.0
			Q17	24073	4.0	Q37	128028	22.4
			Q18	0	0.0	Q38	12803	2.2
Q0	594882	100	Q1	594882	100			
heat exchanger			Q2	570809	100	Q3	570809	100

Table 7.26 Heat balance in the incineration process under scenario 4

	MJ/d	%		MJ/d	%		MJ/d	%
Q01	1097723	36.6	Q11	1237509	41.2	Q31	737697	25.1
Q02	1244693	41.5	Q12	1076181	35.9	Q32	1076181	36.6
Q03	288033	9.6	Q13	553531	18.4	Q33	333358	11.3
Q04	7638	0.3	Q14	62150	2.1	Q34	62150	2.1
Q05	353273	11.8	Q15	6972	0.2	Q35	4335	0.1
Q06	9368	0.3	Q16	4018	0.1	Q36	2498	0.1
			Q17	60367	2.0	Q37	658311	22.4
			Q18	0	0.0	Q38	65831	2.2
Q0	3000727	100	Q1	3000727	100			
heat exchanger			Q2	2940361	100	Q3	2940361	100

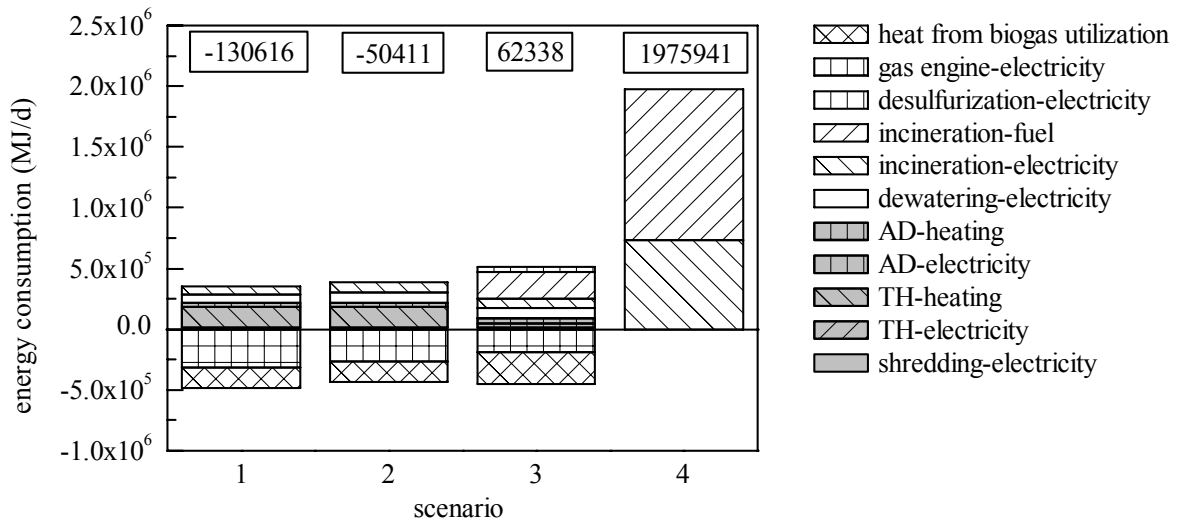


Figure 7.8 Energy consumption

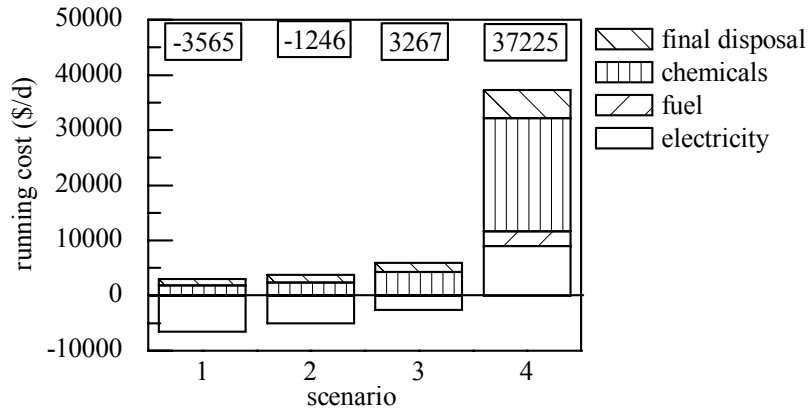


Figure 7.9 Running cost

Scenario 1 and 2, which include TH process, is found economically feasible. As discussed in the former section, energy needed for heating in TH process can be covered by waste heat from incineration and gas engine. As a result, little running cost is needed for this process. However, it greatly improves digestibility and dewaterability. Accordingly, less cost are needed for other processes like incineration and desulfurization, while more money are obtained from sale of electricity produced from biogas. As ASBR improved digestibility and dewaterability further, more profit is produced in scenario 1 than in scenario 2.

Scenario 3 costs 3267 \$/d, but is still much less than scenario 4. As the treatment capacity of incineration in scenario 4 is much larger than other scenarios, the cost gets huge. It implies that treatment of MBW by AD or other processes before final disposal can save cost.

7.7.4 Greenhouse gas emission

Emission of greenhouse gas inside the system is shown as Figure 7.10.

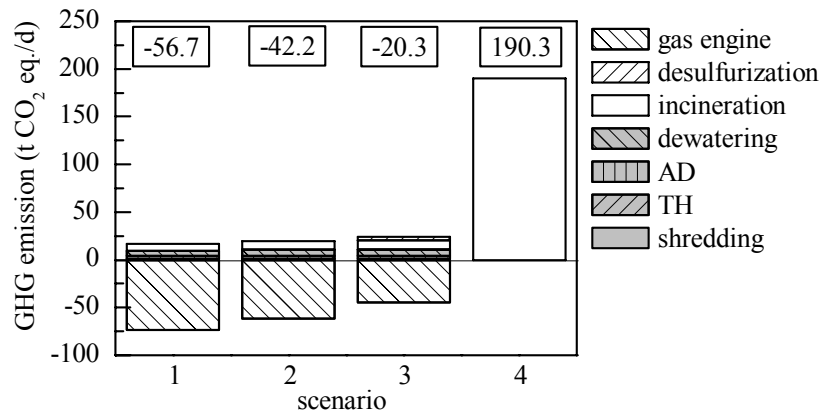


Figure 7.10 Greenhouse gas (GHG) emission

It can be found scenario 1-3 can be environmental friendly with GHG reduction, while scenario 4 produces 190.3 t-CO₂ eq./d of GHG. These differences result from two aspects. Firstly, the treatment capacity of incineration process is much less in scenario 1-3. Secondly, electric output can be obtained in scenario 1-3 by utilizing biogas produced from the AD process. As a result, GHG reduction is possible under these scenarios. Maximum GHG reduction is obtained in scenario 1. With TH and ASBR to improve digestion efficiency and dewaterability, 73.5 t-CO₂ eq./d of GHG can be reduced by outputting electricity from gas engine.

7.8 Conclusion

Energy consumption and GHG emission is scenario 1 < scenario 2 < scenario 3 < scenario 4, while running cost is contrary.

It can be concluded, anaerobic digestion of the MBW before final disposal by incineration can help to reduce energetic, economic, and environmental loads of the MBW treatment system. Both application of thermal hydrolysis and ASBR can help to reduce these loads further, which makes the whole system be energetically, economically, and environmentally friend. It implies promising future to use this TH-ASBR combined technology for the treatment of MBW.

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Chapter 8 Conclusion and suggestions

8.1 Conclusion

In this research, anaerobic digestion of municipal biowaste was studied in a pilot-scale plant in Beijing, China. Thermal hydrolysis (TH) was applied as pre-treatment to improve digestibility of the biowaste. Its effects on properties of the biowaste and digestion performance were then evaluated. Except for improving solid hydrolysis, thermal hydrolysis was found able to improve settling performance of the biowaste. According to this, anaerobic sequencing batch reactor (ASBR), in which a solid settling phase is designed, was used as a digester. Solid settling behaviors in the ASBR were studied and ASBR was compared with continuously stirred tank reactor (CSTR) for digester stability and performance. Furthermore, a model was established to predict organic removal and microorganism growth in both ASBR and CSTR from an importation parameter- solid retention time (SRT). At last, systematic evaluation was made to the TH-ASBR combined technology from the aspects of energy, economy, and environment. All these contents are aimed to provide experiences and suggestion for future studies on this TH-ASBR combined technology for the treatment of municipal biowaste. Results were also compared with lab studies and other studies for verification.

Major conclusions are as following:

- 1) 38.3% of volatile suspended solids (VSS) in the municipal biowaste were hydrolyzed by thermal hydrolysis pre-treatment. Volatile fat acids (VFAs), which are important inhibitors in the digestion process, also increased by 42%. As buffering ability of the biowaste was also improved by thermal hydrolysis, digestibility was not affected by the VFA increase. Biochemical methane potential (BMP) test showed maximum methane production rate was improved by 115%. These results were greater than those obtained in lab studies, as the pilot plant used steam for heating while the lab studies used electricity. The latter may cause unequal heating in the biowaste and thus weaken hydrolysis performance. Thermal hydrolysis also improved fluidity of the biowaste. Viscosity was decreased. Settling and dewatering performance was improved. These resulted from interaction of molecule destruction, particle

size variation, dilution, and so on. Especially, improvement on settling performance indicated that application of ASBR become possible.

2) Pilot-scale digesters with and without TH pre-treatment were compared for their stability and performance under organic loading rates (OLRs) of about 1.5 and 3 kg VS/(m³·d). It's found thermal hydrolysis improved digester stability, especially under a higher OLR of about 3 kg VS/(m³·d). Obvious difference was not found for performance of the two digesters on organic removal and daily biogas production. However, batch monitoring results revealed that reaction in the digester without TH pre-treatment was lagged for about 4-6 h. It implied digestion was accelerated when TH pre-treatment was applied. When digestion time is shortened (under a lower hydraulic retention time (HRT)), digesters without TH may get unstable easier than those with TH pre-treatment.

3) Solid settling behaviors in ASBR were studied. At lower OLRs of 1.6-1.8 kg VS/(m³·d), settling behaviors can be described by combination of zone settling and compression settling. Settling viscosity can be predicted from suspended solids (SS) concentration by Vesilind equation. Viscosity, which was easier to be determined than SS, was found relate to (SS) and able to predict the settling viscosity. However, at a higher OLR, settling performance was interrupted by biogas production and could not be explained by the classical theory. ASBR was found outperforming CSTR in many aspects. It had better stability, lower effluent strength, better removal of organics, and more biogas production than CSTR. Especially, ASBR performance showed better resistance to HRT decrease than CSTR, which indicated digester capacity can be improved when ASBR is applied.

4) A model developed from the hypothesis of hydrolysis limitation was established for prediction of substrate removal and microorganism growth in both ASBR and CSTR. Effluent concentrations and solid accumulation in ASBR can also be predicted from this model. Another well-proved model was also used for comparison. It was found the model established in this thesis corresponded with the experiment data, and were also comparable with the simulation results from the other model. It showed SRTs should be longer than about 20 days to obtain relative stable removal of the organics in MBW, which also corresponded with common suggestion from other studies. This result indicated an advantage of ASBR over

CSTR, as longer SRTs can be obtained in ASBR than CSTR when they are operated under the same HRT.

5) Systematic evaluation was made to the TH-ASBR combined technology from the aspects of energy, economy, and environment. Treatment of municipal biowaste by TH-ASBR before incineration, were compared with TH-CSTR, CSTR, and direct incineration. Results showed the energy consumption and greenhouse gas (GHG) emission is TH-ASBR + incineration < TH-CSTR+ incineration < CSTR + incineration < incineration, while running cost is contrary. It indicated anaerobic digestion of biowaste before final disposal by incineration can help to reduce energetic, economic, and environmental loads of the whole system. Both application of TH and ASBR can help to reduce these loads further, which makes the system be energetically, economically, and environmentally friend. It implies promising future to use this TH-ASBR combined technology for the treatment of MBW.

8.2 Suggestions

8.2.1 From the view of this study

1) Municipal biowaste treated in this study was a mixture of fruit-vegetable waste, restaurant kitchen waste, and wasted sewage sludge in a ratio of 1/1/1. However, studies were not carried out on the most suitable mixing ratio of these wastes. Other wastes were also not taken into consideration. Evaluation on the mixing ratio and waste type are needed in future studies.

2) Thermal hydrolysis in pilot scale showed a little advantage over that in lab scale in Chapter 3. The reason may be different heating source used in these two studies. However, experimental conditions like scale and waste properties were different in these two studies. As a result, further confirmation is needed to verify this advantage.

3) In Chapter 4, batch experiments were carried out for monitoring the variation of VFAs and other parameters from feeding to discharging. But all the batch experiments were carried out for just once. Repeated analyses are needed to verify the results.

4) Solid settling behaviors in ASBR were found obey to classical settling theory under a relative low OLR in Chapter 5. However, solid settling was interrupted by biogas production under higher OLRs and was not able to be described by this theory. As a result, further studies are needed to take the effect of biogas production into consideration.

5) A model based on the hypothesis of hydrolysis limitation was established in Chapter 6. Although the fitting results were acceptable comparing with experimental results and other models, better modeling may be achieved if considering the soluble and particle organics respectively. Moreover, as the amount of microorganisms and solid substrates can't be separately determined in this study, the model cannot give separate description of change of these two parameters. Further studies should separate the microorganism from substrates by other indicators.

6) In Chapter 7, systematic evaluation showed running cost could be reduced when anaerobic digestion is used. But construction cost was not considered during system evaluation, which may affect the results of cost a lot. Some parameters like viscosity can have

large effect on the operating cost. However, they were not considered in this study. In TH process, steam consumption was not considered. It's very different from results of pilot plant. Uncertainty analysis should also be carried out to make more accurate evaluation of the system. Results also need to be compared with data from actual operation of the relative plants.

8.2.2 From the view of municipal biowaste management

TH-ASBR combined technology was found technically, energetically, environmentally, and economically feasible for the treatment of municipal biowaste before final disposal in this study. However, this technology also has its limitation. For some easily-biodegradable biowastes, their digestion rate is fast enough and application of TH may not be required. Some studies even showed TH treatment decreased digestibility of biowaste. As solid settling performance in ASBR is affected by solid concentration and biogas production, ASBR application under high solid concentration and high OLRs may be unrealizable. As a result, this technology is especially suggested for treatment of municipal biowaste which is hard to be hydrolyzed, under relative low solid concentrations and low OLRs.

For situations under other conditions, other technologies/methods are suggested for the management of municipal biowaste. Some biowaste like papers can be recycled. Garden Waste with relative low water content, like straw and garden waste, can be treated by pyrolysis. Waste with high solid concentrations, can be digested in other digesters like two-phase digesters and anaerobic contact digesters; while waste with low solid concentrations can be treated by digesters with mobile phase, like upflow anaerobic sludge bed (UASB), expanded granular sludge bed (EGSB), and digesters with filters or membrane.

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