ANAEROBIC SEQUENCING BATCH REACTOR TREATMENT OF LOW STRENGTH SWINE MANURE AND CO-DIGESTION OF ENERGY DENSE BY-PRODUCTS

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ANAEROBIC SEQUENCING BATCH REACTOR

TREATMENT OF LOW STRENGTH SWINE

MANURE AND CO-DIGESTION OF ENERGY

DENSE BY-PRODUCTS

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Abstract: The Anaerobic Sequencing Batch Reactor (ASBR) is a batch digester utilizing a single vessel for digestion and solids separation. Internal solids retention provides the ASBR with the ability to treat dilute solids wastewaters such as swine manure. A 430 m^3 full scale ASBR at the Oklahoma State University Swine Research and Education Center was started in August of 2008 using a "cold start" technique and continuously operated for two years. After one year of operation, the HRT was incrementally reduced from 20 to 5 days, and the cycles per day was increased from one to two at a temperature of 20°C. Operation at the final parameters provided organic removals of 64% and a specific methane yield of 0.33 m^3 CH_4 kg VS^{-1}. The dilute natural of swine manure provides the opportunity for co-digestion utilizing the ASBR. Co-digestion of crude glycerol from biodiesel production and swine manure in lab-scale ASBR systems was examined to determine the maximum stable crude glycerol inclusion rate for operation at a 5 day HRT and temperature of 20°C. The maximum inclusion rate for these parameters was found to be 1% (v/v) of the daily influent volume. The 1% inclusion resulted in a 7.3 fold increase in methane production, 21.2 l CH_4 day^{-1} compared 2.9 l CH_4 day^{-1} for swine manure only. A 92.5% conversion of crude glycerol COD to methane was observed at the 1% inclusion rate indicating the near complete utilization of co-digestion feedstock. The ASBR has been reviewed in detail lab-scale experiments with limited experience in full scale design and operation. Considerations for the design of a full scale ASBR for treatment of low strength swine manure should be based upon reactor energy balance and solids retention. The low organic loading of low strength swine manure results in low volumetric energy production necessitating the design process to include the reactor's operational input energy requirements with respect to available energy production. The inclusion of the solids separation process within the reactor requires that the physical reactor design be based upon solids settling rates and solids retention capacity of the reactor vessel.

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LIST OF SYMBOLS

B_c	Contois kinetic coefficient, dimensionless
CHP_H	Combined heat and power heat recovery efficiency, %
CHP_E	Combine heat and power electrical energy recovery, %
C_{eff}	Effluent solids concentration, kg m^{-3}
C_{MB}	Biogas methane concentration, m^3 CH_4 m^{-3} biogas
C_{OE}	Volatile solids concentration of decanted effluent, kg m^{-3}
C_{OI}	Influent OM content, kg m^{-3}
C_{OML}	Volatile solids concentration of mixed liquor, kg m^{-3}
C_{OS}	Volatile solids concentration of wasted sludge, kg m^{-3}
CSS	Sludge volatile suspended solids concentration, kg m^{-3}
D	Reactor diameter, m
D_{eff}	Motor drive efficiency, frac.
DIHR	Design influent heating requirement, kWh m^{-3} reactor volume day^{-1}
DRPR	Design reactor transfer pumping requirement, kWh m^{-1} day^{-1}
E_{pump}	Pump energy requirement, kw-hr day^{-1}
Eff_{VSS}	Effluent volatile suspended solids concentration, kg m^{-3}
F	Substrate utilization rate, mass $volume^{-1}$ $time^{-1}$
HRT	Hydraulic retention time, days
H	Required pump pressure head, kPa
H_{dec}	Mixed liqour decant phase height, m
H_{ro}	Reactor operating height, m
H_{romax}	Maximum reactor operating height, m
H_{sd}	Settling distance, m
HMPV	Influent heating required methane production volume, m^3 CH_4 m^{-3} day^{-1}
I_{om}	Influent organic matter concentration, mass $volume^{-1}$
K_s	Monod kinetic half velocity coefficient, mass $volume^{-1}$
k_d	Death rate coefficient, $time^{-1}$
MLS	Mixed liquor solids concentration, kg m^{-3}
MLVSS	Mixed liquor volatile suspended solids concentration, kg m^{-3}
MSC	Microbial substrate conversion efficiency, frac.
OLR	Organic loading late, kg OM m^{-3} day^{-1}
OLR_H	OLR for Influent Heating, kg COD m^{-3} day^{-1}
OLR_P	OLR for reactor transfer pumping, kg COD m^{-3} day^{-1}
ORE	Organic Matter Removal Efficiency, %
P_{eff}	Pump Efficiency, frac.
PMP	Potential methane production, methane volume reactor $volume^{-1}$ $time^{-1}$

PMPV	Transfer pumping methane production volume, m^3 CH_4 m^{-3} day^{-1}
Q	Transfer flow rate, m^3 s^{-1}
Q_B	Volume of biogas produced, m^3 day^{-1}
Q_E	Effluent flow rate, m^3 day^{-1}
Q_I	Influent flow rate, m^3 day^{-1}
R	Cycles per day, day^{-1}
RPR	Reactor transfer pumping requirement, kWh m^{-3} day^{-1}:
S	Concentration of degradable substrate in the effluent, mass $volume^{-1}$
SMLS	Settled mixed liquor solids concentration, kg m^{-3}
SMY	Specific methane yield, m^3 CH_4 kg^{-1} OM
S_o	Concentration of degradable substrate in the influent, mass $volume^{-1}$
SPR	Sludge production rate, kg day^{-1}
SRT	Solids retention time, days
SUR	Microbial substrate utilization rate, kg COD kg VSS^{-1} day^{-1}
t_c	Cycle time, day
t_{comp}	Time required to reach compression settling, time
t_d	Decant phase length, m
t_F	Transfer phase length, hr
t_R	React phase length, day
t_s	Sludge wasting period, day
t_{smax}	Maximum settling time, time
t_{smin}	Minimum settling time, time
t_{SRT}	Time period for SRT calculation, days
t_t	Time required to reach transition settling, time
T_R	Reactor temperature, °C
T_I	Influent temperature, °C
TIHR	Theoretical influent heating requirement kWh m^{-3} reactor volume day^{-1}
v_s	Solids settling velocity, m min^{-1}
V_c	Cycle volume, m^3
V_{ML}	Volume of mixed liquor lost during depth sensor malfunction, m^3
V_{ro}	Reactor operating volume during react phase, m^3
V_s	Volumetric substrate utilization rate, mass $volume^{-1}$ $time^{-1}$
V_{sludge}	Volume of sludge removed, m^3
VOLR	Volumetric organic loading late, kg OM m^{-3} day^{-1}
VRE	Volumetric reactor efficiency, m^3 CH_4 m^{-3} reactor day^{-1}
W_B	Brake pump power, kW
W_P	Pump Power, kW
Y	Maximum cell-yield coefficient as cell mass per substrate mass, mass $mass^{-1}$
ΔM_{MLO}	Change in mixed liquor organic matter, kg
μ_m	Maximum growth rate, $time^{-1}$
Θ	Hydraulic or solids retention time, day
Θ_H	Hydraulic retention time, day
Θ_S	Solids retention time, day

CHAPTER I

INTRODUCTION

Introduction

The Anaerobic Sequencing Batch Reactor (ASBR) is a batch anaerobic digester which includes the digestion and solids separation process steps within a single vessel. The inclusion of the solids separation process step within the reactor allows for the separation of the hydraulic and solids retention times without the need for external clarification and sludge recycle. The performance and operational characteristics of the ASBR are examined in a review of ASBR and anaerobic digestion literature.

Start-up and continuous operation of a full scale ASBR for the treatment of low strength swine was completed. The objective of the start-up of a full scale ASBR was to examine the ability to utilize a cold start for start-up of a full scale reactor treating low strength swine manure. The cold start utilized only the low strength swine manure to seed the reactor rather than the addition of anaerobic digester sludge from an existing reactor. Two objectives were examined during the continuous operation of the full scale digester. First, the examination of low strength swine manure as feedstock for continuous stable operation of a full scale ASBR. Second, the examination of the operational parameters and performance of a full scale ASBR treating low strength swine manure.

The dilute natural of swine manure and the ASBR's treatment capacity for organic loading rates above that provided by swine manure, allows for the opportunity for co-digestion. Co-digestion refers to the inclusion of a secondary organic influent stream to supplement the primary feedstock to increase the reactor's organic loading and biogas production. Crude glycerol an energy dense by-product of biodiesel production was chosen for examination as a co-digestion feedstock for an ASBR treating swine manure. The high energy density of crude glycerol compared to swine manure allows for significant biogas production increases at low inclusion rates. The objective for the examination of crude glycerol as a co-digestion feedstock for swine manure was to determine the maximum inclusion rate for operation at a 5 day HRT and 20°C. This examination includes measurement of the digestibility of crude glycerol when co-digested with swine manure and resulting biogas production increase from crude glycerol inclusion.

The ASBR has been reviewed in detail regarding its operational parameters and potential treatment performance in laboratory scale experiments. In the full scale the ASBR has limited experience with only two full scale reactors having been constructed. The objective of the final chapter is the development of design steps for consideration in development of full scale ASBR's for treatment of low strength swine manure. The design considerations developed to meet the objective are based upon the potential reactor energy balance and solids retention. The low organic loading rate obtained from swine manure results in low volumetric energy production. This necessitates that the design process include the estimation of the reactor's operational input energy requirements. Secondly, the inclusion of the solids separation process within the reactor requires that the physical reactor design be based upon solids settling rates and solids retention.

CHAPTER II

REVIEW OF LITERATURE

Introduction

The Anaerobic Sequencing Batch Reactor (ASBR) is a single vessel batch anaerobic digestion reactor developed and patented at Iowa State University (U.S. Patent No. 5,185,079) (Sung and Dague, 1995). The ASBR operates by cycling through a sequence of four phases in a single reaction vessel; fill, reactor, settle and decant (fig. 1). The inclusion of the settling phase within the reactor vessel provides the ASBR with the ability to separate the hydraulic retention time (HRT) (eq. 1) from the solids retention time (SRT) (eq. 2) without external clarification and sludge recycle (Zhang et al., 1997; Sung and Dague, 1995). The HRT is average time the influent substrate volume is retained within the reactor, while the SRT is the average time sludge is retained within the reactor. The SRT is the most critical of the two parameters as the retention of sludge, microbial biomass, affects the treatment performance and stability of the reactor. The ability to separate the HRT and SRT allows the ASBR to be operated at HRT's less than the desired SRT while maintaining treatment of the influent organic waste stream. The SRT is a function of the HRT for a continuously stirred anaerobic digestion reactor (CSTR) without solids recycle, thus the effluent solids concentration is nearly equal to the mixed liquor solids concentration (Chen and Hashimoto, 1980). To increase the SRT of a CSTR, the HRT must be increased accordingly or include effluent clarification with sludge recycle (Traverso et al., 1988).

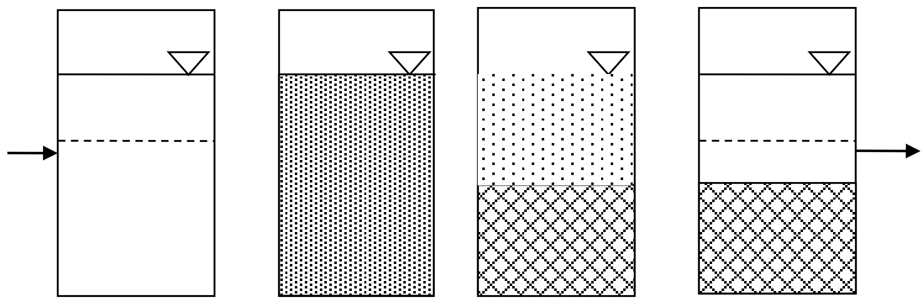

Figure 1. Phases of One ASBR Cycle (Steele and Hamilton, 2009).

The hydraulic retention time for an ASBR is calculated as follows:

$$HRT = V_{ro} / V_c * R \tag{1}$$

Where

HRT = Hydraulic retention time, days

V_c = Cycle volume (m^3)

R = Cycles per day (day^{-1})

V_{ro} = Reactor operating volume during react phase (m^3)

The solids retention time (SRT) is calculated as follows:

$$SRT = (V_{ro} * MLVSS) / (V_c * R * Eff_{VSS} + C_{SS} * V_{Sludge} / t_s) \tag{2}$$

where

SRT = Solids retention time (days)

MLVSS = Mixed liquor volatile suspended solids concentration (kg m^{-3})

Eff_{VSS} = Effluent volatile suspended solids concentration (kg m^{-3})

C_{SS} = Sludge volatile suspended solids concentration (kg m^{-3})

V_{Sludge} = Volume of sludge removed (m^3)

t_s = Sludge wasting period (day)

The ASBR's ability to retain solids and separate the HRT and SRT is demonstrated by Wang et al. (2009). An ASBR and CSTR model reactor were operated in parallel treating thermally hydrolyzed sewage sludge. During the startup of the reactors, both models were operated as CSTR's until day 80. After day 80 the hydraulic flow operation of one of the models was changed to ASBR, which in figure 2 is marked by the significant reduction in effluent total solids (TS) and the start of reactor solids accumulation. As the reactor TS concentration continued to increase, the SRT of the ASBR reactor, at day 10 HRT, averaged 37 days.

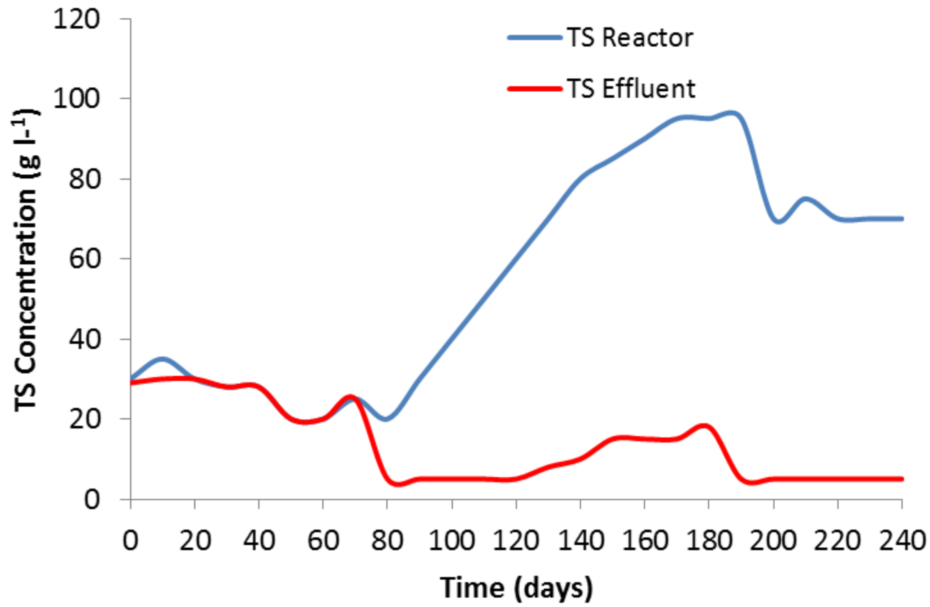

Figure 2. Effluent and reactor total solids concentration for model ASBR treating thermally hydrolyzed sewage sludge with respect to time adapted from Wang et al, 2009

The ASBR reactor in figure 2 also had a marked increase in the removal of organics compared to the parallel CSTR reactor. The Volatile Solids (VS) and total Chemical Oxygen Demand removals were 10% and 20% higher for the ASBR during the 20 and 10 day HRT operational periods. What should be noted is the soluble Chemical Oxygen Demand (SCOD) removal for the both the ASBR and CSTR at both 10 and 20 day HRT's were approximately equal, with values between 91.6 and 92.6%, respectively. This data shows that increase in VS and TCOD removal is a direct result of the ASBR solids retention. Biogas production of ASBR increased 15% and 30% at 20 and 10 day HRT's, respectively, compared to the CSTR, indicating that the settled solids were not just stored but digested (fig. 3). Similarly, Hansen et al., 1999, found significant improvement in reactor performance by including a settling period prior to effluent withdrawal in a model CSTR. A 52% increase in gas production was obtained by the addition of the settling phase along with a 33% and 27% increase in reactor TS and VS, respectively.

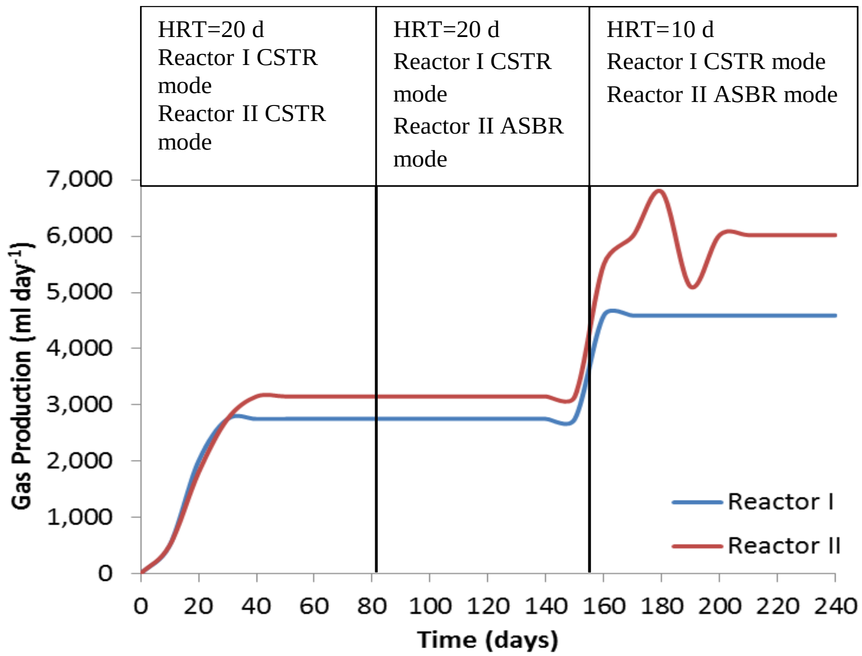

Figure 3. Daily biogas production for ASBR and CSTR model reactors treating thermally hydrolyzed sewage sludge adapted from Wang et al., 2009

ASBR Operational Parameters

The operational parameters controlling the potential performance of an anaerobic digester are; HRT, SRT, temperature, volumetric loading rate (VOLR, g COD l^{-1} day^{-1}), and specific organic loading rate (SOLR, g COD g VSS^{-1} day^{-1}). For the CSTR the relationship of these parameters is related to influent concentration and reactor HRT. As the ASBR is able to separate the HRT from the SRT through internal solids separation the operational parameters are affected by the solids settling and solids retention. As the SRT increases in the ASBR the mixed liquor solids concentration is increased reducing the SOLR as it is a function of mixed liquor volatile suspended solids concentration.

With increased SRT and mixed liquor volatile solids concentration the resulting reduced SOLR can provide the ability for reduced reactor operating temperatures (Dague et al., 1998). The relationship and impact of these operational parameters on the ASBR will be discussed in further detail.

Anaerobic Digestion Process

Hydrolysis of Organic Solids

The speciation of solids for wastewater divides the substrate solids into seven divisions (fig. 4). Total Solids mass is the residual mass remaining after the evaporation of the liquid substrate at 103 to 105˚ C (APHA, 1998). The dissolved solids are the solids fraction that is passed through a filter with a nominal pore size 0.45 μm and suspended solids is the fraction that is retained by the filter. The fixed and volatile solids differentiate the material between inorganic and organic. Fixed solids remain after ignition of the dried solids residue at 550˚ C.

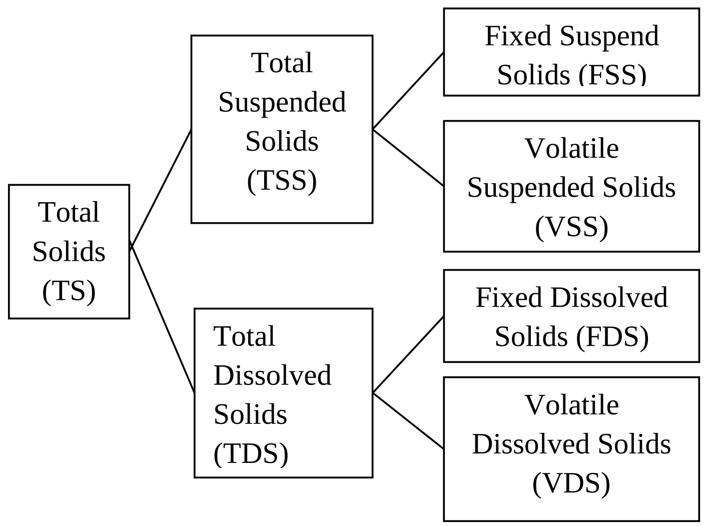

Figure 4. Gravimetric solids fractions.

During anaerobic digestion VSS and VDS are converted to organic acids by acid forming bacteria which are then converted to methane and carbon dioxide by methane forming bacteria (fig. 5) (McCarty, 1964). In an ideal anaerobic digestion system the effluent would contain only FDS and FSS as these are undigestible. The ideal scenario is first limited by gravimetric solid separation which cannot differentiate between fixed and volatile suspended solids. Secondly, 100% of the volatile solids (organic) fraction is not biodegradable under anaerobic conditions for example the lignin fraction of plant biomass (Ghosh and Christopher, 1985).

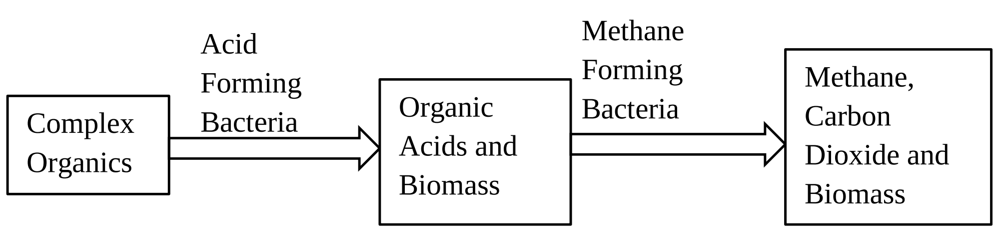

Figure 5. Organic substrate conversion during anaerobic digestion (McCarty, 1964)

Applying the ideal anaerobic digestion effluent scenario to the ASBR the first two considerations are the HRT and the cycle react phase length. The react phase must be adequate and practical for the removal of the soluble volatile solids fraction. Figure 6 shows the substrate concentration during multiple cycles of an ASBR (Sung and Dague, 1995). The F/M ratio referred to in figure 6 is the feed to microbial mass ratio similar to the SOLR (g COD g VSS^{-1} day^{-1}). The F/M ratio and the SOLR are a function of both the HRT, feed mass, and SRT, microbial mass, which are independently controlled in the ASBR. Achieving a low substrate concentration and F/M at the end of the cycle time provides optimal effluent quality by reducing the soluble volatile solids concentration and reduced biogas evolution aiding flocculation and settling of biomass (Sung and Dague, 1995).

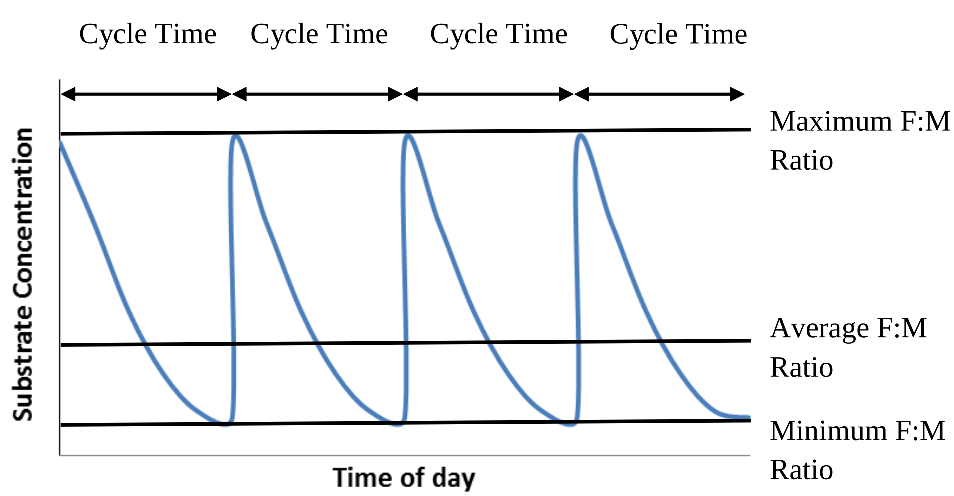

Figure 6. Feed to microbial mass ratio during operation of ASBR adapted from Sung and Dague, 1995

Hydraulic Retention Time

The HRT is a measure of the time the liquid fraction of the influent waste stream is maintained within in the reactor assuming no short circuiting. The HRT is expressed in time (t) which may be minutes, hours or days depending upon the operation of the reactor.

Table 1 lists HRT's of published laboratory scale ASBR experiments indicating the potential wide range of ASBR HRT's. For full-scale operation only two reactors have been constructed and operated both for the treatment of swine manure. The ASBR constructed and operated by Iowa State University researchers was operated at a HRT of 15 days (Angenent et. al, 2002). The second ASBR was constructed and operated at Oklahoma State University (OSU). The OSU ASBR was initially operated at an HRT of 20 days and later reduced to 5 days with no significant change in reactor performance (Steele and Hamilton, 2009 and 2010).

Table 1. ASBR hydraulic retention times, operational temperatures and influent substrates citied in literature

Hydraulic Retention Time	Operational Temperature	Influent Substrate	Source
5, 2.5, and 1.25 days	35°C	Synthetic glucose wastewater	Cheong and Hansen, 1999
24, 16, 12, 8, and 6 hours	5, 10, 15, 20 and 25 °C	Nonfat Dry Milk	Dague et al., 1998
3.3, 5, and 10 days	35 and 55°C	Activated sludge	Hur et al., 1999
12, 6, 8, and 4 days	20 and 35°C	Swine manure	Ndegwa et al., 2005
4 days	20 and 35°C	Swine manure	Ndegwa et al., 2008
48, 24, 16, and 12 hours	15, 20, 25, and 35°C	Nonfat Dry Milk	Ndon and Dague, 1997
48, 24, and 12 hours	35°C	Nonfat Dry Milk	Sung and Dague, 1995
10, 8, 6.6, 5, 3.3, 2, 1.5 days	35°C	Landfill leachate	Timur and Ozturk, 1999
20 and 10 days	35°C	Thermally hydrolyzed sewage sludge	Wang et al., 2009
24 hours	33°C	Brewery wastewater	Xiangen et al., 1999
6, 3, and 2 days	25°C	Swine manure	Zhang et al., 1997

The optimum HRT for the ASBR based upon the results of the studies listed in table 1 is a function of the influent substrate. Ndegwa et. al., 2005 found that for dilute swine manure the optimum HRT's were 5.25 and 6 day for operation at 20 and 35°C based upon specific gas yield 0.14 ml biogas mg COD^{-1} and 0.16 ml biogas mg COD^{-1}, respectively. Zhang et. al. 1997 also treated swine manure and the data provided indicating an optimum HRT of 6 days compared to 2 and 3 days at a temperature of 25°C. For a low solids synthetic wastewater produced from nonfat dry milk the results of Sung and Dague, 1995 indicated that for VOLR's (eq. 3) between 2 and 8 g COD l^{-1} day^{-1} there was no difference in specific gas yield for HRT's of 12, 24 and 48 hours. Likewise, Cheong and Hansen, 1999 found little difference in gas production with respect to HRT for ASBR's fed a low solids glucose based synthetic wastewater. The solids characteristics of the influent substrate will impact the HRT. As the suspended solids fraction increases the HRT must increase to allow for the solids particles to be hydrolyzed. This is evident when comparing the soluble low solids synthetic substrates with HRT's of less than 48 hours to the 5 to 6 day HRT required for swine manure.

The volumetric organic loading rate (VOLR) is calculated as follows:

$$VOLR = (Q_i * I_{om}) / V_{ro} \quad (3)$$

Where

VOLR = Volumetric organic loading rate, (kg OM m^{-3} day^{-1})

Q_i = Influent flow rate, (m^3 day^{-1})

I_{om} = Influent organic matter concentration (mass $volume^{-1}$)

Solids Retention Time

The SRT for anaerobic digestion refers to the average time that the microbial biomass is retained within the reactor. The SRT is the ratio of the mass of the microbial biomass in the system to the mass leaving the system with respect to time. Typically it is assumed that the microbial biomass is equivalent to the Volatile Suspended Solids (VSS) fraction of the mixed liquor and treated effluent. SRT calculation for an ASBR is given eq. 2.

Figure 7 highlights the separation of the HRT and SRT of 58 laboratory scale ASBR's. SRT values of 1, 4 and 6 times the reactor HRT are indicated by solids lines. The ASBR's independence of HRT and SRT through the retention of solids via internal solids separation is illustrated by an average SRT:HRT ratio of 12 with a median value of 7.5 for these reactors (figure 8). The average and median SRT are 28 and 20 days with average and median HRT's of 3.9 and 2.5 days. To achieve similar SRT's, a CSTR reactor volume would be 12 times that of similarly operated ASBR. No specific optimum SRT:HRT ratio or SRT length was determined or discussed in the model ASBR studies. However, the average and median SRT length of these studies do follow the basic guidelines for anaerobic digestion. Metcalf and Eddy(2003) state that at a treatment temperature of 30°C a minimum of 20 days is needed and Dague (1981) suggests that a SRT of 15 days is required to achieve VS removals of 50%.

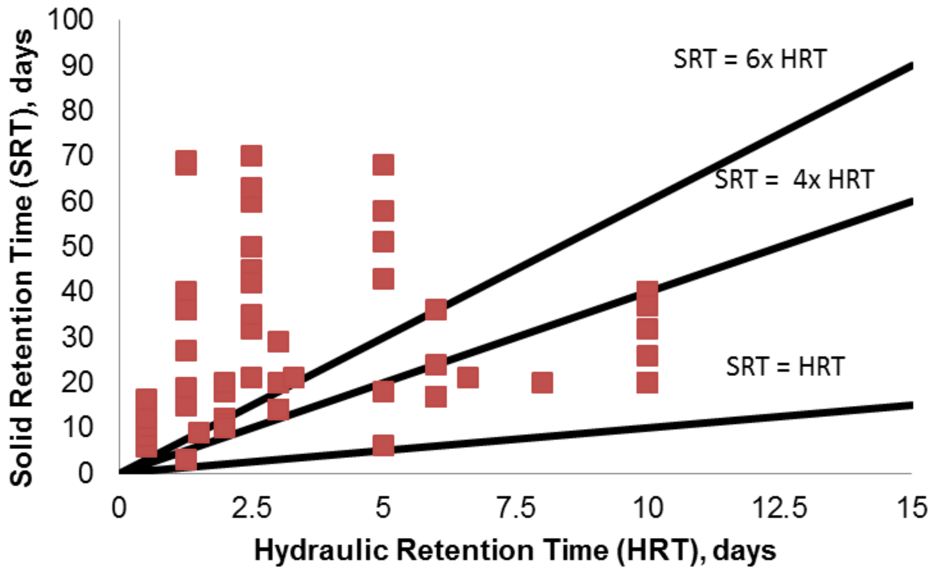

Figure 7. ASBR solids retention times with regard to HRT (data from Cheong and Hansen, 2008; Zang et al., 1997, Wang et al., 2009; Timur and Ozturk, 1999; and Lee et al., 2008)

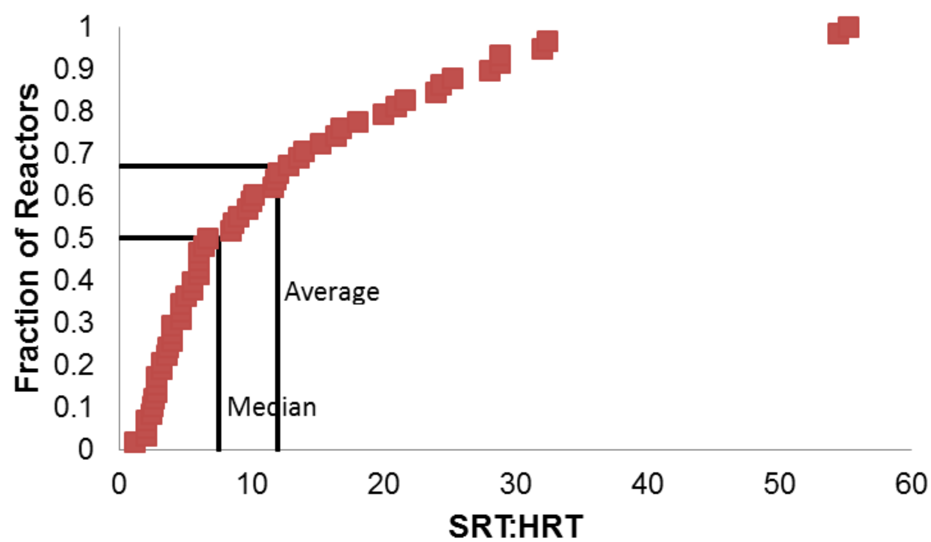

Figure 8. Distribution of ASBR SRT:HRT ratios (data from Cheong and Hansen, 2008; Zang et al., 1997, Wang et al., 2009; Timur and Ozturk, 1999; and Lee et al., 2008)

Settling

Five gravitational sedimentation phenomena can be observed in an anaerobic digester; discrete particle settling, flocculent settling, hindered settling, compression and flotation (Metcalf and Eddy, 2003). The last phenomenon, flotation, is normally not desired as this would result in the washing out of microbial biomass and undigested organic particulates. The type of settling that occurs is a function of the mixed liquor solids concentration.

Discrete particle settling occurs at low solids concentrations where solids settle as individual particles and settling velocities are equivalent to the particle's terminal velocity. During flocculent settling suspended particles flocculate during sedimentation and the increase in particle mass increases the settling rate of the particles. This phenomenon also occurs in solutions of low solids concentrations. As the solids concentration increases the settling regime changes from discrete and flocculent to hindered and compression. Compression settling occurs at high solids concentrations where the particles form a structure and continued settling of the particles is a result of the increasing mass of the particle structure. This type of settling is found in the lower layers of settled solids where the solids are allowed to remain undisturbed.

Hindered settling also called zone settling occurs in systems of low to high solids concentrations. In these systems compression, discrete and flocculent settling may occur during sedimentation. As a result of increased solids concentrations the particles are in contact and do not act as single particles. The mass of particles settle together maintaining a relative position in the particle mass forming a zone or blanket. As the zone layer moves downward a clear supernatant layer is formed above the zone layer. As the zone layer solids concentration increases the settling regime transitions from hindered to compression settling.

The suspended solids concentration in the reactor not only determines the settling phenomena but the settling rate or velocity. Ndegwa et al., 2001, examined the effect of suspended solids concentration on the sedimentation of undigested swine manure. Although the substrate reviewed in Ndegwa et al. (2001) has not been anaerobically digested the general trend is applicable. Figure 9 shows the difference in TSS removal with regard to time and concentration.

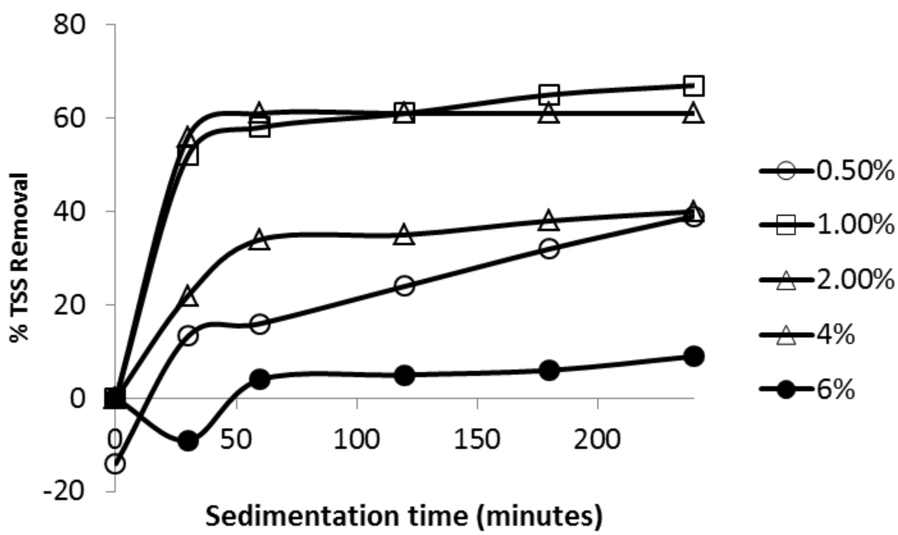

Figure 9. Total suspended solids removals from undigested settled swine manure with respect to total solids concentration adapted from Ndegwa et al., 2001

Plotting the data shown in figure 9 with TSS removal as a function of concentration the removal trend is more pronounced (figure 10). The trend in figure 10 shows that there is an optimal total solids concentration range for TSS removal via settling; between 1% and 2% total solids for this substrate. Although figure 10 is based upon undigested swine manure the solids removal curve shape is similar to that which would be produced during a hindered settling solids flux analysis for activated sludge (Metcalf and Eddy, 2003). This similarity provides the basis for the assumption of similar settling velocity trends for anaerobically digested fecal materials. The trend of reduced settling velocity with increase solids concentration was also measured and observed by Sung and Dague (1995). In this study a synthetic wastewater of nonfat dry milk was fed to four equal volume ASBR models with varying depth to diameter ratios. The narrowest reactor achieved a settling velocity of approximately 9 cm min^{-1} at a mixed liquor suspended solids concentration of 10 g l^{-1} and the other three reactors a velocity of 5 to 6 cm/min. As the mixed liquor suspended solids concentration increased up to 30 g/l the difference in settling velocities was reduced to a range of 0.2 to 1.8 cm/min.

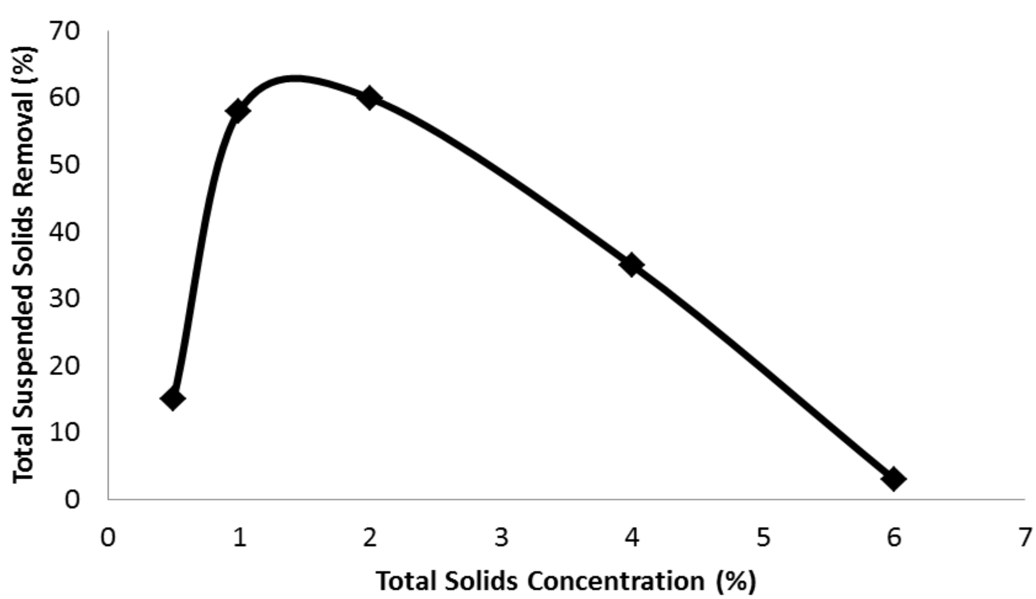

Figure 10. Total suspended solids removal after 60 minutes settling time (Ndegwa et al., 2001)

Sludge Wasting

The internal solids separation of ASBR requires the additional process of sludge wasting to remove excess solids from the reactor. The removal of excess solids for management of the mixed liquor solid concentration provides the ability to maintain a consistent solid settling velocity within the reactor. As shown figure 10 and discussed in Ndegwa et al., 2001; there is an optimal mixed liquor solids concentration with regard to solids settling a given influent substrate. If solids are allowed to continuously accumulate a mixed liquor solids concentration will be reached at which time settling is hindered; resulting in increased effluent solids concentration and biomass washout.

As the mixed liquor solids concentration increased in the model ASBR described by Wang et al., 2009 the effluent solids concentration began to increase reducing the reactor solids accumulation rate. This is shown in figure 2 beginning around day 140 when the mixed liquor solids concentration reached 60 g l^{-1}. Upon reaching a mixed liquor solids concentration of 90 to 95 g l^{-1} biogas production began to drop from 6.79 to 5.10 l day^{-1}. At this point 600 ml of sludge was removed from the reactor on day 186 and effluent quality and biogas production returned to previous levels. Based upon the solids accumulation rate it was determined that 300 ml of sludge was to be removed every 10 days to maintain a constant mixed liquor solids concentration, effluent quality, and biogas production rate.

Mixing

Experiments studying slurry mixing provide a wide variety of data and correlations. Mixing intensity measurements are based on several mixing indicators;

complete of the bottom mixing, slurry height, and vertical solids concentration profile. Typically complete off the bottom mixing and slurry height are used in determining mixing power requirements. Often the mixed liquor solids concentration is still stratified when only using these parameters.

Determination of required mixing intensity requires some experimental data for prediction of homogeneity and power requirements. This is especially true for the dynamic characteristics of manure slurries. The prediction of the mixing intensity required for complete mixing or partial suspension of solids cannot be universally predicted. To predict complete homogeneous suspension of solids within the reactor individual experimentation of the desired mixing regime and slurry are required. As seen in figure 11 the mixing intensities provided or calculated from the literature are highly variable ranging from a USEPA recommended intensity of 6.6 W m^{-3} to 3,500 W m^{-3} utilized by Bhutada and Pangarkar, 1989., with no specific justification for the intensity selection for model reactor experiments.

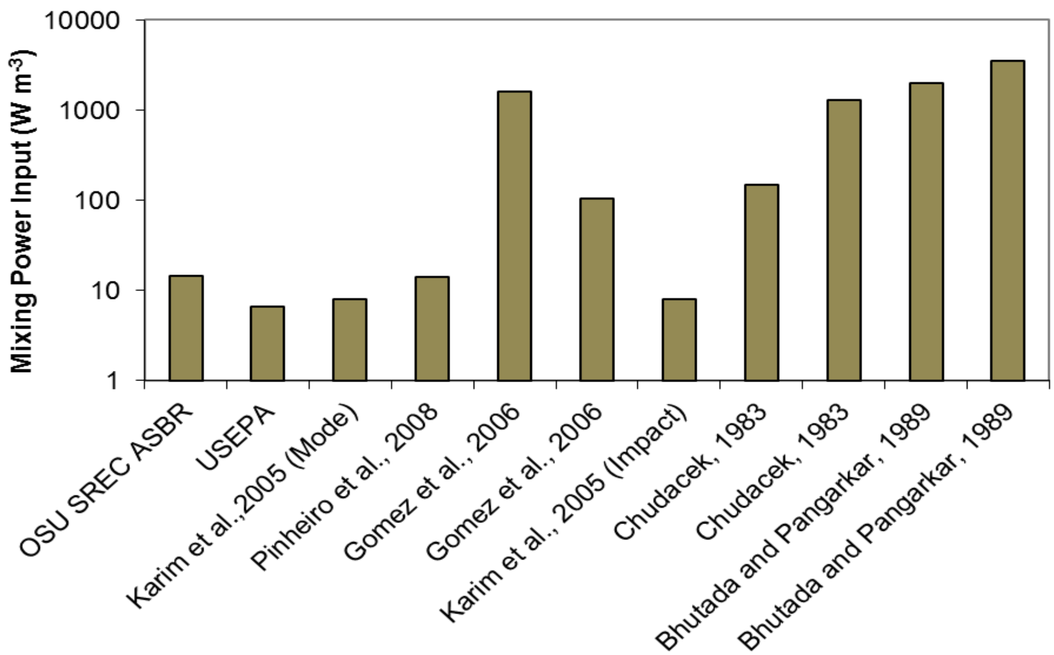

Figure11. Mixing intensities from literature for anaerobic digestion reactors

There are three typical modes of mixing in anaerobic digestion; mechanical, mixed liquor recirculation, and biogas recirculation (USEPA, 1979). Regardless of mixing mode the objectives of mixing are; temperature maintenance, substrate distribution, sedimentation prevention, scum and crust prevention, and release of entrapped gases (Mills, 1979; Ward et al, 2008). Each of the three mixing modes is capable meeting these needs however; there are limitations and drawbacks for each mode. For liquor recirculation and biogas recirculation high solids concentrations limit the ability of mixing and result in clogged nozzles (Karim et al., 2005). Additionally, the use of biogas recirculation for mixed liquor mixing has the potential for the introduction of air into the system during the compression and return of the biogas to the reactor.

The solids concentration, reactivity of the substrate, and reactor geometry must be considered when selecting a mixing system for the ASBR. The geometry of the tank can create dead zones on the floor of the reactor and corners in rectangular reactor designs. For mechanical mixing the design process with regard to tank geometry, tank structure, and operational liquid level become more significant than use liquor recirculation, i.e. jet mixing (Bathija, 1982). Mechanical mixers require additional structural support for both the exterior power unit and interior impeller. The mechanical components of a mechanical mixing system require additional maintenance, methane as it is a paraffin, breaks down the grease in bearings and seals (Mills,1979).

The literature does not indicate an optimum mixing intensity based upon reactor performance. Based upon the USEPA recommendation and Karim et. al., 2005, the initial design intensity should be 6 and 8 W/m^3. The results of Karim et. al. (2005) show that the concentration of the mixed liquor solids determine mixing need and intensity. At solids concentration less than 5% TS, Karim et al. (2005) found no significant difference between the three modes of mixing and unmixed reactors. At TS concentration of 10%, mixing increased biogas production by 22% compared to unmixed reactors. Gomez et al, 2006 showed no difference in biogas production between mixing intensities of 105 and 1,600 W/m3. With mixed liquor solids concentrations of less than 2% this follows the same trend found by Karim et. al. (2005). The optimum mixing intensity for any anaerobic digestion reactor is the one that provides temperature maintenance, substrate distribution, sedimentation prevention, scum and crust prevention, and release of entrapped gases with the lowest input energy requirement.

All three modes of mixing have been employed in of model and pilot scale ASBR's and two in full scale. The Iowa State University full scale ASBR in Nevada, IA constructed by Iowa State University utilized two 3 kW propeller mixers (Angenet et al., 2002). The Oklahoma State University ASBR located in Stillwater, OK utilizes a three nozzle jet mixer located on the bottom of the reactor (Steele and Hamilton, 2009; Steele and Hamilton, 2010). For ASBR's with low mixed liquor solids concentrations treating dilute substrate mixing mode and intensity are not significant in the performance of the reactor compared to operation at solids concentrations above 5 g/l (Karim et. al., 2005). When treating low strength waste this is increasingly important as the volumetric gas production rate is less than for high strength wastes thus the volumetric energy input into the reactor must be reduced to produce a net energy gain. This was observed by Steele and Hamilton (2010) for the Oklahoma State University ASBR. The design flow rate for the jet mixing system was set at 69 l/s (1,100 gpm) and was reduced to 9.5 l/s without negative impact on reactor operational performance. This reduction reduced the daily power requirements from 253 to 129 kW-hr/day.

Temperature

The ASBR's ability to retain and accumulate an active biomass within the reactor provides the ability for the system to compensate for reduced biological rates at temperatures less than 35˚C (Dague et al.,1998). ASBR operating temperatures in the literature range from 5 to 55˚C as shown in in table 1. The retention and accumulation of active biomass allows for COD removal of 75% and higher for operating temperatures in the mesospheric and psychrophilic range (Dague et al., 1998; Ndegwa et al., 2005;

Ndegwa et al., 2008). Dague et al. 1998, found that at a HRT of 6 hours and an operating temperature of 5˚C soluble COD and Biological Oxygen Demand removals of 65 and75% were achieved and 90% removals at 25˚ C when treating nonfat dry milk synthetic wastewater. Methane production for the 6 hour HRT 5 and 25 °C treatments with an OLR of 2.4 g COD l^{-1} day^{-1} resulted in specific methane yields 0.1 and 0.26 l CH_4 g COD^{-1} day^{-1}, respectively (Dague et al., 1998). In two studies conducted at Oklahoma State University it was found that for low strength swine manure the ASBR performed better at a temperature of 20˚ C than at 35˚ C (Ndegwa et al., 2005 and Ndegwa et al., 2008).

With the ability to operate at temperatures less than 35˚C significant input energy reduction is available. When considering ASBR operation for a swine facility or industrial wastewater stream from an indoor source the influent temperature is close to room temperature thus greatly reducing the need for heating of the influent prior to feeding. The heating requirements are then a factor of the heat loss from the reactor during operation when ambient temperatures are less than the operating temperature. The energy savings to operate an ASBR at a temperature of 20˚C compared to 35˚C is 224,880 BTU (65.9 kWh) per 1,000 gal. In terms of methane this a savings of approximately 225 ft^3 (6.37 m^3) of methane per 1,000 gallons. Assuming a natural gas cost of $8 per 1,000 ft^3 this equates to energy savings of $657 per year per 1,000 gallons.

Digestion Substrates

For an anaerobically digestible organic substrate to be utilized as a feed stock in an ASBR three characteristics must be considered; alkalinity, pH, nutrient content, and settling. The optimal pH range for anaerobic digestion is between 6.6 and 7.6 (McCarty, 1964). For maintenance of the pH within this near neutral range the recommended alkalinity should be with 2,000 and 5,000 mg / l as $CaCO_3$ (Metcalf and Eddy, 2003; McCarty, 1964). As with any biological process macro and micro nutrients are required for microbial activity. Nutrient requirements for nitrogen, phosphorus, and sulfur based upon microbial biomass composition range from 10 to 13, 2 to 2.6 and 1 to 2 mg 100 mg of biomass^{-1}, respectively (Metcalf and Eddy, 2003). For maximum activity the mixed liquor soluble nutrient concentration for nitrogen, phosphorus, and sulfur should be 50, 10, and 5 mg l^{-1}.

The carbon to nitrogen ratio, C:N, can also impact the biogas production from a given substrate during anaerobic digestion. Hills (1979) reported on digested diary manure that had been amended with glucose and cellulose to augment the C:N ratio. Figure 12 shows the resulting change in reactor gas production with regard to C:N ratio, indicating that the optimal C:N ratio is near 25. The reduction of biogas production at low C:N ratios is attributed to ammonia inhibition and nitrogen limitations at high C:N (Llabres-Luengo and Mata-Alvarez, 1988).

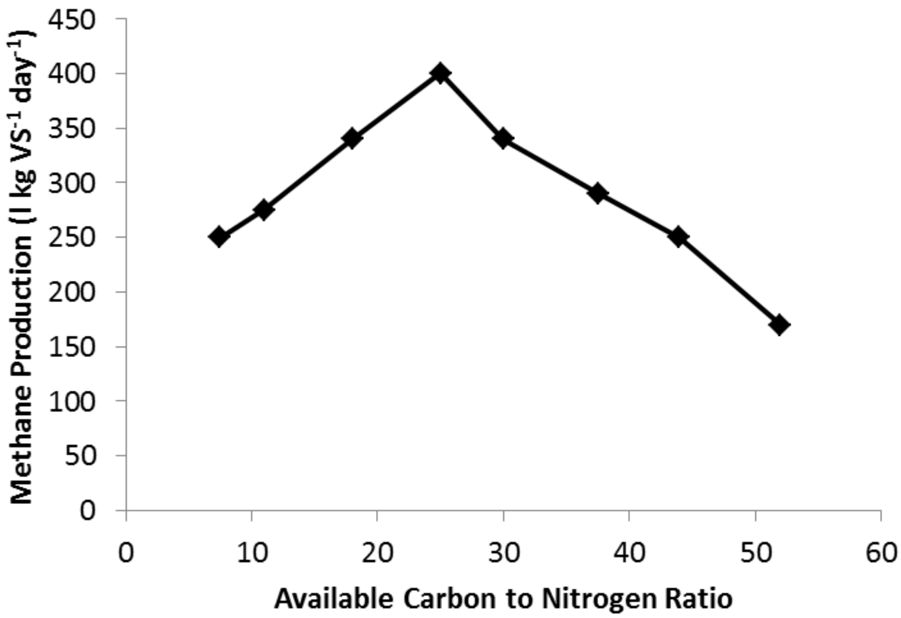

Figure 12. Methane production rate with respect to carbon to nitrogen ratio adapted from Hills, 1979

Anaerobic digestion has been applied for the treatment of organic waste streams other than those from fecal sources such as manure and municipal and domestic wastewaters (table 2). The studies listed in table 2 treated nontraditional substrates and wastewaters using anaerobic digestion. Several of the studies utilized a fecal waste stream and an additional organic substrate which is referred to as co-digestion. The use of co-digestion provides not only an additional means for treatment of an organic waste stream but also the opportunity for improved reactor performance. As described above the balancing of the C:N ratio of the influent substrate provides increased biogas production. Co-digestion amendment substrates with high C:N ratios provide the ability to increase the C:N ratio of livestock manures which typically have low C:N ratios

Table 2. Anaerobic co-digestion and non-fecal based substrates

Substrates	Source
Water hyacinth, coastal Bermuda grass and processed municipal solid waste	Ghosh and Christopher, 1985
Castor cake	Gollakota and Meher, 1988
Fruit and vegetable fraction of municipal solids waste and primary sludge	Gomez et al., 2006
Tomato solid wastes (peels, stems, and seeds)	Hills and Nakano, 1984
Landfill leachate	Kennedy and Lentz, 2000
Hemp thermomechanical pulping wastewater	Kortekass et al., 1998
Glycerol from biodiesel production	Lopez et al., 2009
Potato processing wastewater and glycerol	Ma et al., 2008
Dairy whey	Mockaitis et al., 2006
Sisal fiber waste	Mshandete et al., 2006
Personal care industry wastewater	Oliveria et al., 2009
Cattle manure and Rice Straw	Pathak et al., 1985
Boreal herbaceous grasses	Seppala et al., 2009
Municipal solid waste and biosolids	Stroot et al., 2001
Landfill leachate	Timur and Ozturk, 1999
Brewery wastewater	Xiangwen et al., 2008
Organic fraction of municipal solid waste and biosolids	Zhang et al., 2008

To maintain the ASBR’s ability to effectively separate the HRT and SRT the solids settling must be maintained. The settling velocity of the co-digestion substrate must be such that when combined with the mixed liquor no significant change in settling velocity it observed. Additionally, for particulate co-digestion substrates the increased influent solids concentration will resulted in increased mixed liquor concentrations resulting in reduced zone settling velocities and extended settling phase length. The increased in OLR will also increase the solids accumulation rate due increased solids addition and biomass production requiring adjustment of the sludge wasting period.

Kinetics

The use of kinetic models for design and operation of anaerobic treatment systems provides a method for predicting reactor stability and performance. Both the Monod and Contois models have been used in the modeling of biological waste treatment systems (Smith, 1981; Chen and Hashimoto, 1980; Dague et al., 1998). The Monod model for microbial kinetics assumes that effluent substrate concentration is independent of the influent substrate concentration, whereas the Contois model assumes dependence (Smith, 1981).

The Monod model effluent substrate concentration for a CSTR is calculated as follows:

$$S = K_s*(\Theta_H*k_d + 1) / [\Theta_H*(\mu_m - k_d) - 1] \qquad (4)$$

where

S=concentration of degradable substrate in the effluent (mass volume^{-1})

K_s=Monod kinetic half velocity coefficient (mass volume^{-1})

Θ_H=Hydraulic retention time (time)

k_d=death rate coefficient (time^{-1})

μ_m=maximum growth rate (time^{-1})

k_d=death rate coefficient ($time^{-1}$)

The Contios model effluent substrate concentration is calculated as follows:

$$S = B_c*Y*(S_o – S) / [\Theta_H*(\mu_m – k_d) – 1] \quad (4)$$

where

B_c=Contois kinetic coefficient (dimensionless)

Y=maximum cell-yield coefficient as cell mass/substrate mass (mass $mass^{-1}$)

S_o= concentration of degradable substrate in the influent (mass $volume^{-1}$)

Chen and Hashimoto(1980) reported that the Monod model was not suited for anaerobic digestion of complex organic wastes in part due to its independence of the effluent and influent substrate concentration. Utilizing the Contois model they derived the substrate utilization rate for a completely mixed, continuous flow anaerobic digestion system without solids recycle (HRT = SRT) (eq. 6).

$$F = (S_o / \Theta)[1-K / (\mu_m*\Theta – 1 + K)] \quad (6)$$

where

F = substrate utilization rate (mass $volume^{-1}$ $time^{-1}$)

S_o= influent substrate concentration (mass $volume^{-1}$)

Θ_H = hydraulic retention time (time)

μ_m = maximum specific growth rate ($time^{-1}$)

K = kinetic parameter (dimensionless)

This relationship for the volumetric substrate utilization rate is the base for the relationship used to express the volumetric methane production rate (Chen and Hashimoto, 1980 and Hashimoto, 1983). By including the ultimate methane yield for the substrate, B_o (volume of methane per mass of substrate), eq. 6 becomes the expression for the volumetric methane yield (eq. 7).

$$V_s = (B_o * S_o / \Theta)[1 - K / (\mu_m * \Theta - 1 + K)] \tag{7}$$

where

V_s= volumetric substrate utilization rate (mass volume^{-1} time^{-1})

S_o= influent substrate concentration (mass volume^{-1})

Θ = hydraulic or solids retention time (time)

μ_m = maximum specific growth rate (time^{-1})

K = kinetic parameter (dimensionless)

B_o = Ultimate methane yield (volume methane mass substrate^{-1})

It should be noted that for the above relationships (substrate utilization and methane production) it was presented in the literature that the HRT and SRT are interchangeable as there was no solids recycle. As reactor design for the substrate utilization rate assumes continuous flow, no solids recycle and complete mixing, this is a valid assumption. However, for an ASBR and other types of reactor designs such as fixed film reactors the SRT and HRT are independent; not assumed to be equal. Thus modification of the substrate utilization and methane production expressions must be modified for use with the ASBR. The volumetric methane production expression can be broken down into two parts; the ultimate methane production volume and the microbial substrate conversion efficiency. The ultimate methane production (PMP) for a given mass of substrate is equal to:

$$PMP = (B_o * S_o / \Theta_H) \tag{8}$$

where

PMP = Ultimate methane production (methane volume reactor volume^{-1} time^{-1})

The microbial substrate conversion efficiency (MSC) at given temperature and influent substrate concentration is:

$$MSC = 1\text{-}K / (\mu_m * \Theta^S - 1 + K) \qquad (9)$$

where

MSC = microbial substrate conversion efficiency (fraction)

Θ_S= Solids retention time (day)

The reactor temperature and influent substrate concentration are used for the estimation of μ_m and K using empirical equations as follows for swine manure (Hashimoto, 1983):

$$\mu_m = 0.013\ (T) - 0.129 \qquad (10)$$

where

T=reactor temperature (°C)

$$K = 0.50+0.0043\ \exp(0.091S_o) \qquad (11)$$

where

S_o= influent substrate concentration (kg volatile solids / m^3)

The for anaerobic digestion systems in which the HRT and SRT are not equal the volumetric methane production (V_s) is calculated as follows:

$$V_s = (B_o\ S_o / \Theta_H)[1\text{-}K / (\mu_m\Theta_S - 1 + K)] \qquad (12)$$

where

V_s = volumetric substrate utilization rate (mass/volume/time)

S_o= influent substrate concentration (mass/volume)

Θ_H = hydraulic retention time (time)

Θ_S =solids retention time (time)

μ_m = maximum specific growth rate of microorganism (1/time)

K = kinetic parameter

B_o = substrate ultimate methane yield (volume methane / mass substrate)

Equation 13 allows for Chen and Hashimoto's expression for volumetric methane production to be utilized for an ASBR by separately identifying the HRT and SRT.

Using the Contois relationships (eq. 5) from Chen and Hashimoto,1980 and Hashimoto, 1983 (eq. 7) alone do not fully account for the ASBR's ability to operate at temperatures less than 35°C. Microbial substrate conversion efficiency for swine manure (eq. 9) is plotted in figure 13 for an ASBR with a 5 day HRT and CSTR with a 30 day HRT both operating at an OLR of 2 g COD/l-day and 30 day SRT. For both reactor types the revised Chen and Hashimoto expression for volumetric methane production (eq. 12) indicates that for temperatures less than 12.5 ˚C no microbial substrate conversion occurs. The maximum specific growth rate (eq. 9) provides the source for temperature inclusion in this relationship. Dague et al. (1998) observed that at a reactor operated at a 1 day HRT with a 25 day SRT the methane production at 5°C was at least 50% of theoretical; which according to the relationship used by Chen and Hashimoto is not achieved until a temperature of 14˚C. Based upon this relationship an increase of operating temperature from 20 to 35˚C should result in a 10% increase in methane production. This was not observed in by Ndegwa et. al., 2008. No significant difference was founds between these two operating temperatures for an ASBR treating dilute swine manure at a 4 day HRT with an approximate SRT of 30 days.

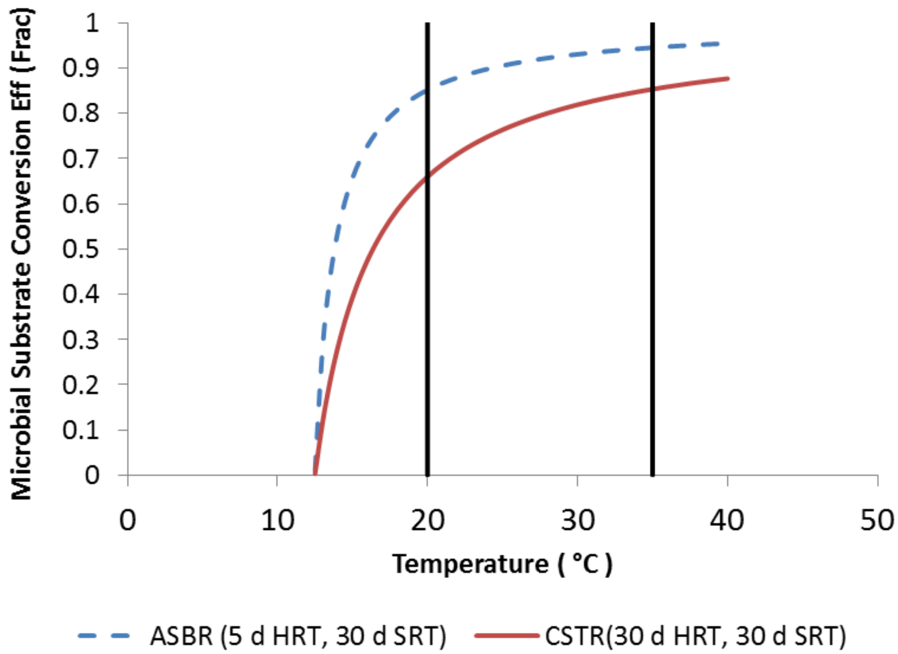

Figure 13. Calculated microbial substrate conversion efficiency (eq. 10) for swine manure with respect to temperature for a 5 day HRT ASBR and 30 day HRT CSTR both with 30 d SRT

Using the HRT and SRT separated expression and K and μ_m relationships from Hashimoto, 1983 the volumetric methane production rate was estimated for 106 trials from 6 studies. The predicted and published volumetric methane production rates are provided in figures 14 and 15. The results of figure 15 show that for all four influent feed stocks (nonfat dry milk, glucose, swine manure, and landfill leachate) the relationship between the actual and predicted methane production rate follows the same general linear trend slightly above a 1:1 line shown. This indicates that the K is dependent only upon influent concentration rather than the makeup as the empirical calculation for K used from Chen and Hashimoto (1983) was determined for swine manure. The linear regression in figure 16 indicates that the expression will under predict the volumetric methane production which is more favorable than over prediction for preliminary design purposes.

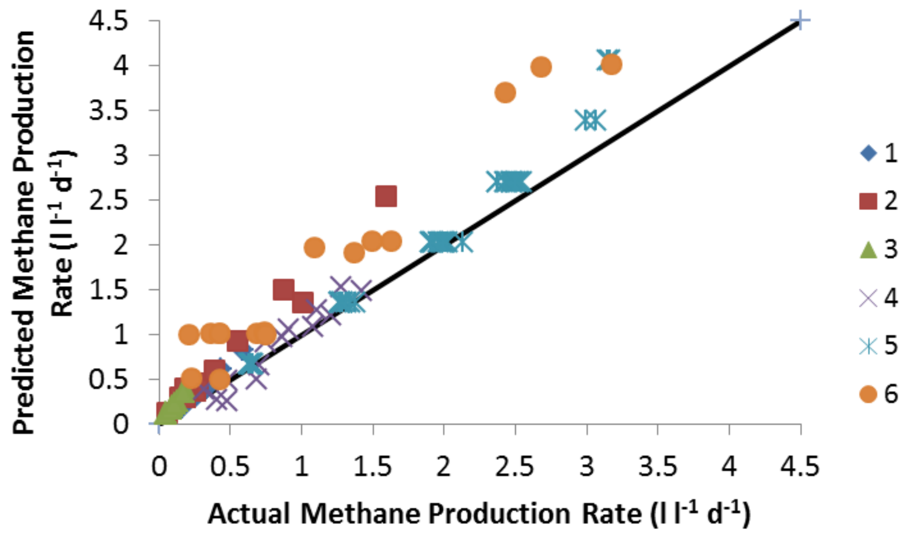

Figure14. Comparison of actual methane production rates and methane production rates predicted by the modified Chen and Hashimoto equation (eq. 12) for multiple feed stocks in laboratory scale ASBR's based upon(1 – Dague et al., 1998: non-fat dry milk;2 – Timur and Ozturk, 1999: landfill leachate; 3 – Ndegwa et al., 2005: swine manure; 4 – Zhang et al., 1997: swine manure; 5 – Sung and Dague, 1995:non-fat dry milk; 6 – Cheong and Hansen, 2008: glucose)

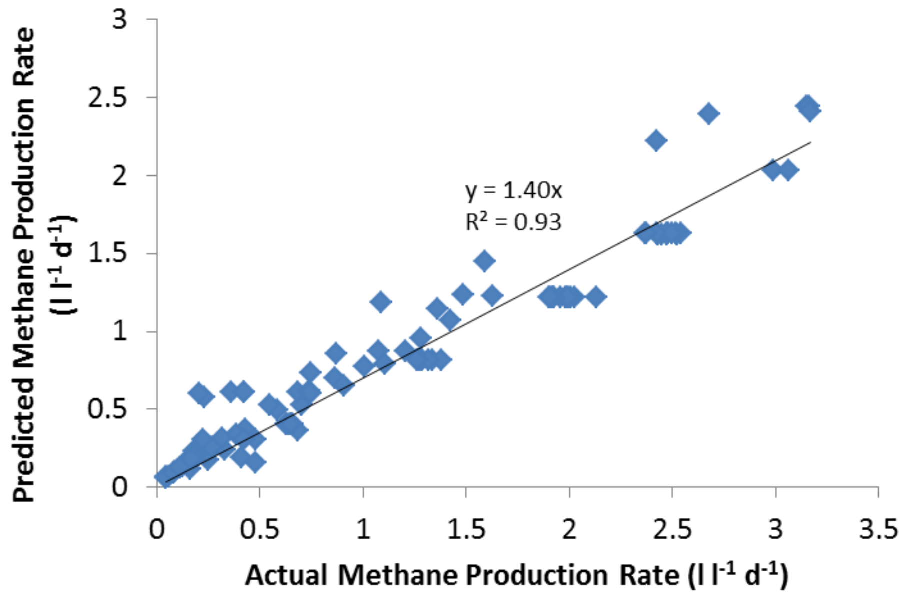

Figure 15. Linear Regression of predicted and measured volumetric methane production rates for laboratory scale ASBR's

Organic Loading Rate

The cyclic F:M ratio shown in figure 6 highlights two F:M ratios the maximum and minimum. The maximum occurring at end of the feed phase and minimum occurring at the onset of the decant phase. The F:M ratio is the ratio of the mass of the COD or VS fed during the feed phase to the mass of mixed liquor VSS also referred to as the SOLR. The mass of the COD or VS remaining at the end of the react phase of an ASBR cycle is a function of the MLVSS, microbial substrate utilization rate, react phase length. Thus for complete substrate utilization the SOLR must be equal to or less than the product of the microbial substrate utilization rate, MLVSS mass, and react phase length as expressed in eq. 13.

$$SOLR \leq SUR \times MLVSS \times V_{ro} \times t_R \qquad (13)$$

where

SUR= microbial substrate utilization rate (kg COD kg VSS^{-1} day^{-1})

MLVSS= mixed liquor volatile solids concentration (kg VSS m^{-3})

t_R= react phase length (day)

Based upon laboratory scale ASBR's experiments the SOLR is approximately 81% of the SUR (figure 16).

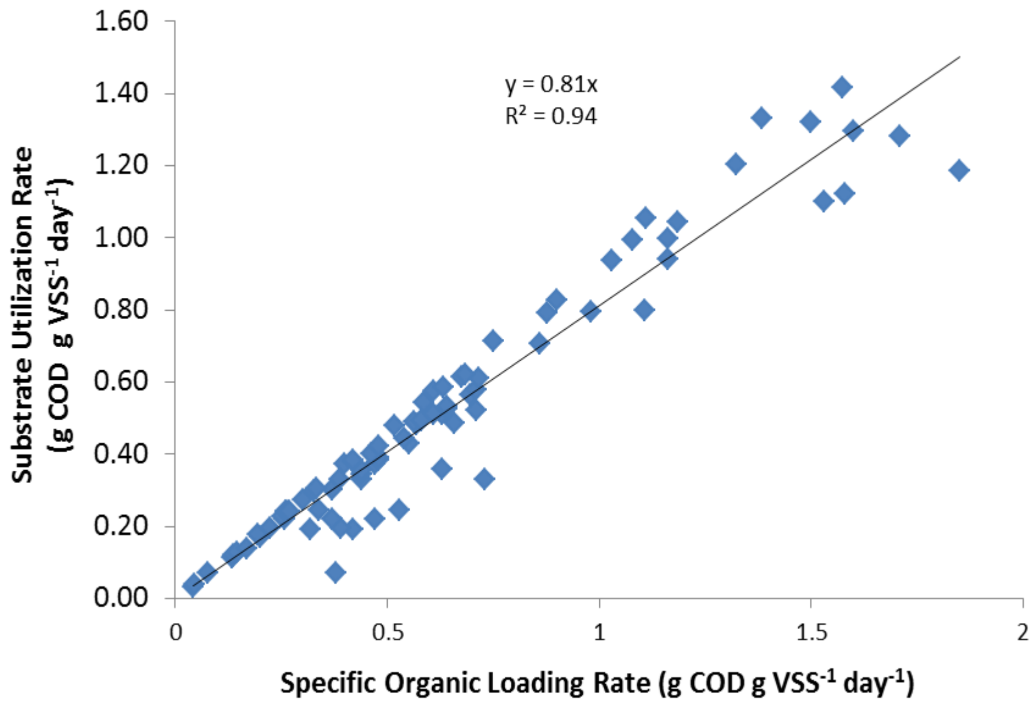

Figure 16. Relationship between Specific Organic Loading Rate and Substrate Utilization Rate of laboratory scale ASBR's (Cheong and Hansen, 2008;Mockaitis et al., 2006; Ndon and Dague, 1997; and Timur and Ozturk, 1999)

Based upon the SOLR:SUR ratio of 0.81 the maximum OLR can be estimated based upon the reactor's maximum mixed liquor suspended volatile solids concentration. As discussed previously the solids settling velocity is a function of the mixed liquor solids concentration. Given there is a maximum mixed liquor solids concentration for optimum solids retention via settling, the maximum mixed liquor solids concentration will determine the maximum optimum OLR for an individual influent waste stream. This further validates the significance of the solids settling and retention in the performance of the ASBR.

ASBR Optimization for treatment of low strength swine manure

Researchers at Iowa State University adapted the ASBR system to operate on swine wastes. (Zhang et al., 1997). Ndegwa, et al. (2005, 2007) working with laboratory ASBR models at Oklahoma State University (OSU) determined the envelope of operating parameters for ASBR digesters treating dilute swine slurries. These laboratory studies provide a baseline for comparison to a full scale ASBR. The laboratory studies examined reactor operational parameters: temperature, HRT, and cycle frequency at 20 and 35° C. The results of the HRT optimization study examined HRT's of 4, 6, 8, and 12 days and found that optimal biogas production occurred at HRT's of 5.25 and 6 days at temperature of 20 and 35°C, respectively. Both studies showed increased COD reduction for the 20°C compared to 35°C; with reduction of 86% to 90% reduction for 20°C and 70 to 86% for 35°C. Effluent volatile fatty acid concentrations (VFA) at 20°C remained lower and more constant than at 35°C. Reactor performance with regard to effluent quality is increased with operation at the lower temperature however specific biogas production is still higher at 35°C. This difference in effluent quality between temperatures is the result of improved settling in the reactors operated at 20°C, however the increased temperature still provides increased microbial activity resulting in a slightly higher specific biogas yield, 0.15 ml mg COD^{-1} compared to 0.14. Reactor COD and VFA reductions, biogas production and specific biogas production for these two studies is provided in table 3 for both temperatures at 1 and 3 cycles per day and HRT's of 4 and 6 days.

Table 3. Model ASBR performance parameters (Ndegwa 2002 and 2005).

Trial Parameters (Temperature, Cycles per day, HRT)	COD Reduction (%)	VFA Reduction (%)	Effluent VFA Conc. (mg l^{-1})	Effluent TSS Conc. (mg l^{-1})	Biogas Production (ml d^{-1})	Specific Biogas Yield (ml mg COD^{-1})
20°C 1 4 d HRT*	90		70	250	2,100	0.146
20°C 3 4 d HRT*	85		70	300	1,700	0.118
35°C 1 4 d HRT*	80		110	400	2,100	0.146
35°C 3 4 d HRT *	70		120	575	1,600	0.111
20°C 1 4 d HRT**	88	86			1,971	0.143
20°C 1 6 d HRT**	89	88			1,442	0.143
35°C 1 4 d HRT**	81	82			2,150	0.152
35°C 1 6 d HRT**	83	86			1,525	0.157

* Ndegwa et al., 2002

** Ndegwa et al., 2005

Conclusions

The diversity of the laboratory scale experiments illustrates the expansive functionality of the ASBR at varying organic loading rates, operational temperatures, mixing regimes, and influent substrates. This diversity in design and application is a result of the ASBR's internal solids separation and solids retention. The cyclic batch process and inclusion of solids separation within that reactor vessel provides the ability to independently control the HRT and SRT. The independent control of the HRT and SRT provides the ASBR with the ability maintain high SRT's and treatment efficiencies at reduced HRT's, 5 days or less.

The ability to control the retention of microbial biomass via internal solids settling and sludge wasting allows for the diverse organic loading rates and operating temperatures of the ASBR. The management of the SRT provides the microbial biomass and particulate substrate retention necessary for effective biological treatment at temperatures less than 35°C. Through active microbial biomass retention low specific organic loading rates are maintained increasing the stability of the ASBR at high volumetric loading rates and reduces recovery times from shock loadings.

Internal solids retention is a distinguishing and diversifying characteristic of the ASBR and principal operational and design parameter for an effective and stable performance. Solids settling provides the mechanism for accumulation and control of the mixed liquor solids concentration which the settling velocity is a function of. The resulting design and operation mixed liquor solids concentration and activity microbial biomass mass controls the design SOLR and VOLR. For the ASBR the solid settling velocity is the single most significant parameter controlling operation and design.

CHAPTER III

START-UP AND CONTINUOUS OPERATION OF A FULL SCALE ANAEROBIC SEQUENCING BATCH REACTOR (ASBR) TREATING LOW STRENGTH SWINE MANURE

Introduction

A 430 m^3 Anaerobic Sequencing Batch Reactor (ASBR) was operated continuously for nearly two years at the Swine Research and Education Center at Oklahoma State University (OSU-SREC). ASBR digesters provide excellent organic matter reduction and efficient energy generation, particularly for low strength wastes. The ASBR is a batch type anaerobic reactor which operates by cycling through a sequence of four phases: fill, react, settle, and decant (figure 1). All four of the phases occur in a single reactor, allowing for settled solids to remain in the reactor. Retention of solids within the reactor allows the hydraulic retention time (HRT) and solids retention time (SRT) to be controlled separately via periodic sludge removal, providing efficient treatment of low strength influents (Sung and Dague, 1995). Low strength influent can refer to low solids concentration, low organic matter content, or both. In either situation, digestion is aided by the ASBR's separation of the HRT and SRT. Adding influent with low organic matter content to a reactor results in a low organic loading rate, and in general, these influents require less time for treatment and benefit from the ASBR's separation of HRT and SRT.

The reduction in reaction time means that effective organic matter reduction can occur at a relatively short HRT. For influent with low solids concentration, retaining solids within the reactor allows a sufficient mass of solids to accumulate in the reactor, providing a surface upon which a biofilm containing methanogens can grow. As HRT is reduced, the ratio of influent to reactor volume increases, and without internal solids retention, microbial biomass may wash out of the reactor. Since solids are retained during the settling and decanting phases of the ASBR cycle, shorter HRT's with a large through-put of liquids are possible without washout of microbial biomass. Reducing reactor HRT reduces reactor volume, which in turn, reduces construction cost. Reducing reactor volume also reduces its heating requirement as the surface area of the reactor is smaller, allowing less heat to escape to the environment. Additionally, reduction in reactor volume increases the reactor's volumetric biogas production efficiency (gas production/reactor volume-time) compared to larger reactors treating the same waste stream.

Objectives

- Examination of the use of a "cold start" technique for an ASBR treating low strength swine manure
- Examination of low strength manure as a feedstock for continuous stable operation of a full scale ASBR

- Examination of the operation parameters and performance of a full scale ASBR treating low strength swine manure

Materials and Methods

Systems Monitored

The full scale ASBR operated at the OSU-SREC is one treatment component in an integrated manure handling and treatment system (figure 17). The farm is essentially a 120 sow farrow to finish operation (Hamilton, et al., 2010). During the time of this study, overall population ranged from 800 to 1,200 animals. Swine are housed in 12 modular style buildings. Each building has a slightly different arrangement, but in general, manure is stored in pull plug pits modified with scrapers. Pits are filled with effluent recycled from the second, aerobic cell of a two-stage lagoon. There are a total of 21 pits on the farm with varying volumes. Manure flows through a 0.20 m diameter sewer line to a splitter box, where it can be directed to either the anaerobic-aerobic lagoon or the ASBR. The average daily manure flush volume is 20.3 m^3 (5,400 gal); flush manure and lagoon effluent characteristics are given in tables 4 and57. Since the 21 small recharge pits are not emptied on a continuous daily basis, a 167 m^3 stainless steel equalization tank buffers the varying daily manure flow.

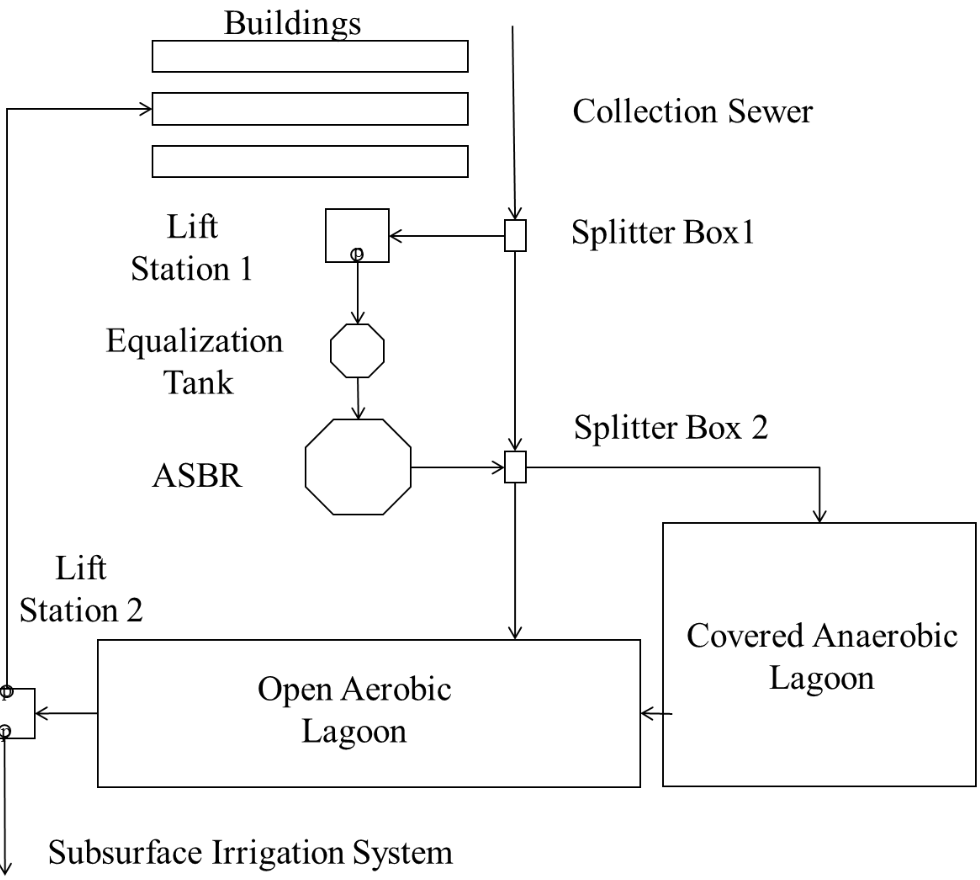

Figure 17. Schematic of OSU SREC manure handling and treatment system

Table 4. Pit Flush Manure Characteristics.

Operational Period	pH			TS			VS			COD		
					mg l^{-1}	mg l^{-1}		mg l^{-1}	mg l^{-1}		mg l^{-1}	mg l^{-1}
	n	X	SD	n	X	SD	n	X	SD	n	X	SD
1	5	6.76	0.3	5	11,400	4,200	5	8,500	3,400	5	15,600	6,300
2	17	6.85	0.2	19	9,400	3,000	19	6,800	2,500	16	13,700	4,500
3a	1	7.0	-	3	8,600	600	3	5,800	500	-	-	-
3b	2	6.72	-	3	9,000	2,300	3	4,900	1,600	1	14,500	-
3c	1	6.67	-	4	9,800	3,100	4	6,900	2,700	1	15,700	-
3d	-	-	-	6	7,300	3,200	6	5,100	2,500	-	-	-
4	-	-	-	3	3,900	900	3	2,400	700	-	-	-

Table 5. Lagoon effluent characteristics.

	8/10/09	9/22/09	11/6/09	1/25/10	X	SD
pH	5.34	5.81	6.14	6.71	6.0	0.6
TS (mg l^{-1})	1,298	647	557	500	750	370
VS (mg l^{-1})	354	389	303	287	333	47
COD (mg l^{-1})	340	140	260	300	260	86

The ASBR itself is an un-insulated concrete tank with a total volume of 430 m^3 (figure 18). The ASBR has 12.2 m (40 ft.) diameter and a unsupported flexible membrane cover (30 mil XR-5 8130, Seaman Corporation). The ASBR has three interior effluent withdraw standpipes allowing changes in HRT, mixed liquor volume, and settling depth. An exterior standpipe houses the reactor depth sensor and serves as emergency overflow. The feed and decant phase lengths are not set by time but by reactor volume set points measured by an ultrasonic depth sensor (EchoSonic II Model LU27-01, Flowline) located in the exterior standpipe (figure 18). Mixing of the ASBR and equalization tank is accomplished using a three nozzle jet mixer located on the floor of each unit. A Fairbanks-Morse B5423 11.2 kW centrifugal pump with 69 liters/s flow capacity at 12.9 m head provides primary mixing for both the ASBR and equalization tank (P1, figure 18). An identical pump is used feed the ASBR.

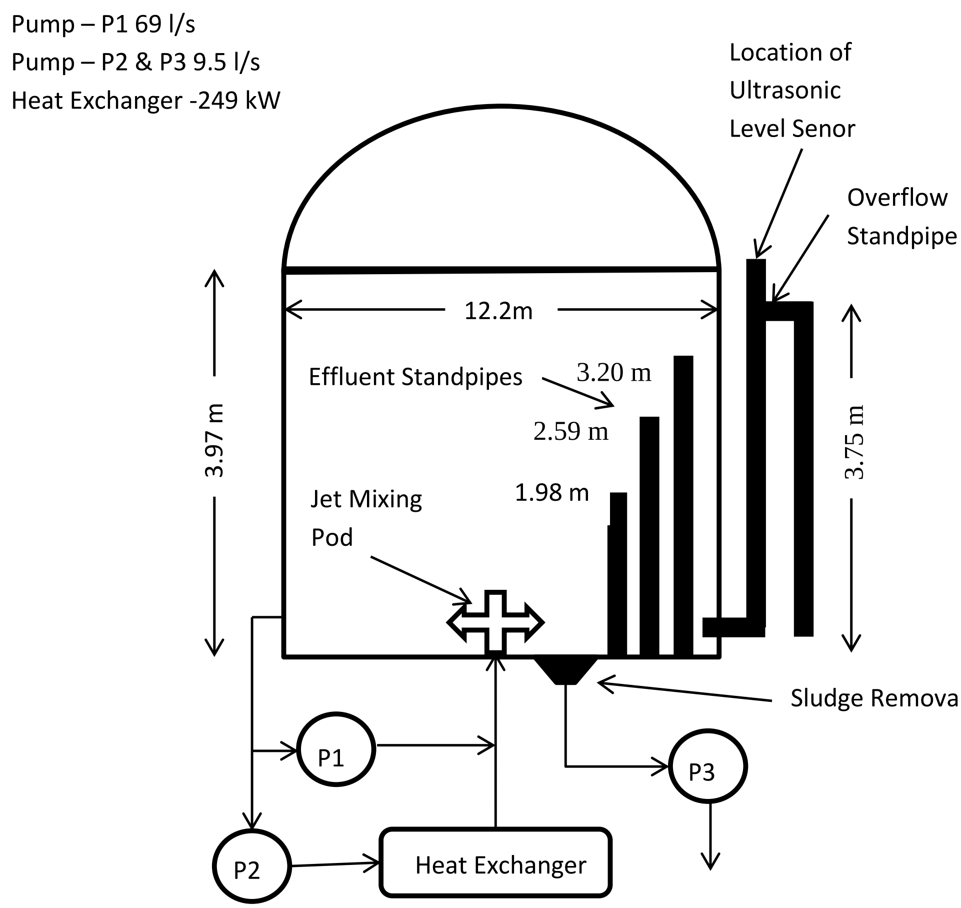

Figure 18. Schematic of ASBR operated at the OSU SREC

Supplementary ASBR heating is provided by an external heat exchanger and natural gas boiler. The Walker Process Equipment HEATX EB boiler is rated for 293 kW and the heat exchanger is rated at 249 kW with a 2.42 m head loss at 9.5 l s^{-1}. Mixed liquor is supplied to the heat exchanger at a flow rate of 9.5 l s^{-1} (150 gpm) and returned via the jet mixing pod located on the bottom of the ASBR(Hayward-Gordon TORUS R2(7), 5.6 kW, P3, figure 18). The heat exchanger return flow adds an additional 0.028 W m^{-3} of mixing. This flow plus the primary mixing flow results in 16.8 ASBR volume turnovers per day, and a mixing intensity of 14 W m^{-3} within the reactor.

The heat exchanger and boiler are only operated when it was necessary to maintain reactor temperature above 20^{o}C; therefore, the reactor is only artificially heated during the time period of December to March. An upper temperature for the ASBR was not maintained. During summer months, reactor temperature could range as high as 37^{o}C due to solar radiation on the black membrane cover and conduction of ambient temperature through the concrete walls of the reactor.

Sludge removal from the bottom of the reactor is provided by a 5.6 kW recessed impeller pump with a flow capacity of 9.4 l s^{-1} located in the ASBR control building (Hayward-Gordon TORUS R2(7), 5.6 kW, P2, figure 18). Biogas produced during the study period was flared.   Biogas production was measured using a Roots Meter Model 15C175 rotary displacement meter. The ASBR pump, influent and effluent valves and biogas system are controlled by an Allen Bradley Panel View550 PLC

Digester Operation

During this study, the ASBR was operated in 8 distinct periods. Start and stop date for each period, as well as operating conditions during the period are given in table 6.

Table 6. ASBR Operational Periods.

Period	Description	Start	End	HRT	Cycles per day	Decant Height	Operating Height	Lagoon Effluent Added	OLR
				days		m	m	$m^3\ day^{-1}$	$kg\ VSm^{-3}\ day^{-1}$
0	Start-up	7/23/08	12/19/08	23.5	1	3.2	3.47	0	0.28
1	Restart and Settling Depth Adjustment	2/2/09	2/28/09	20	1	3.2	3.47	0	0.42
2	Mixing Intensity Adjustment	3/1/09	7/17/09	20	1	3.2	3.47	0	0.37
3a	HRT Reduction	7/18/09	8/9/09	20 -17	1	3.2-2.59	3.47-2.90	0	0.33
3b		8/10/09	9/13/09	16 - 15	2	2.59	2.72	0	0.42
3c		9/14/09	10/11/09	15 -10	2	2.59	2.72-2.76	0 – 24	0.59
3d		10/12/09	12/7/09	10 - 5	2	2.59	2.77-2.92	24 – 48	0.49
4	Steady State	12/8/09	6/7/10	5	2	2.59	2.92	48	0.50

Start-up using "cold start" technique

The OSU SREC ASBR was "cold started" at the end of July 2008. A cold start does not use seed sludge to inoculate the digester, but rather utilizes the natural flora within the influent manure to populate the reactor. Initially, the reactor was filled with water to its maximum depth to support the deflated membrane cover. During start-up this water was replaced with freshly flushed manure. A total of 210 m^3 of manure were added to the reactor over a two week period, 30 m^3 every other day. Biogas production was observed on day 15. Following the observation of biogas production, the ASBR was placed under automated control at an HRT of 23.5 days, one cycle per day, and an OLR of 0.27 kg VS m^{-3} day^{-1} (0.46 kg COD m^{-3} day^{-1}). Biogas production of 79 m^3 day^{-1} (2,800 ft^3 day^{-1}) was measured 28 days after initiation of the cold start, and remained at this level through December of 2008.

Restart and Settling Depth Adjustment.

The ASBR was taken offline in mid-December 2008. It was restarted and operated automatically with one cycle per day on February 2, 2009. Daily influent addition was 20.3 m^3, providing a 20 day HRT and an OLR of 0.42 kg VS m^{-3} day^{-1}. The initial settling phase length was set at 45 minutes based upon previous laboratory work (Ndegwa et al., 2007). During the first several months of continuous operation, mixed liquor samples were collected to measure zone settling velocities of the full-sized reactor. Zone settling velocities were measured by placing well mixed liquor samples into a 61 cm tall glass tube with a 5 cm diameter. The addition of the liquid was marked as time 0 and the time and height of the zone settling solids' interfaced was marked and recorded until solids compression was observed. The zone settling velocity data for the OSU

SREC ASBR mixed liquor followed the expected trend of decreasing velocities with increasing solids concentrations as shown in figure 19. The measurement of the zone settling velocity of the mixed liquor allowed for the determination of the maximum settling distance for the ASBR. Maximum settling distance was set at 0.6 m with a settling time of 60 minutes to achieve this distance based upon an average settling velocity of 2 cm min^{-1} and a factor of safety of 2. Maximum operating depth was set 3.47 m, 0.27 m above the standpipe, and the 60 minute settling time allowed solids to settle 0.33 m below the standpipe before decanting began.

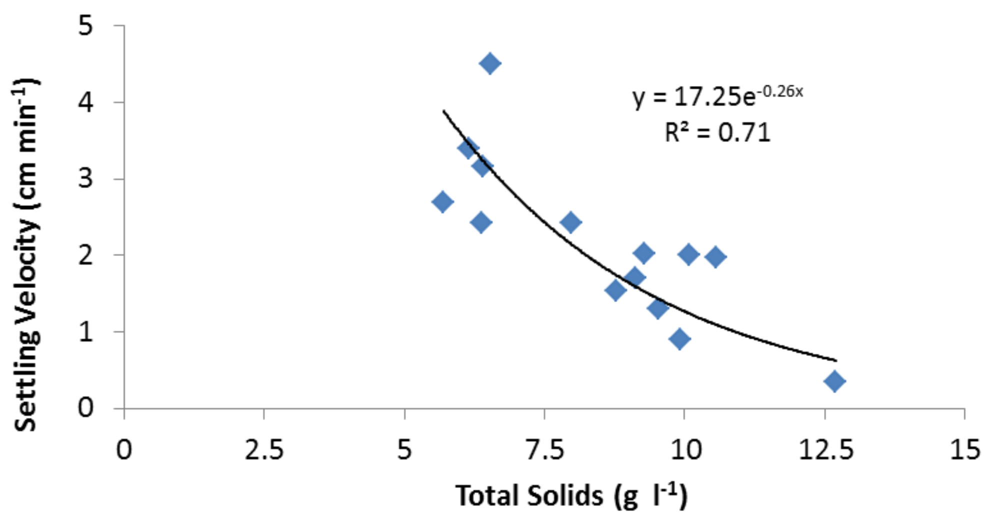

Figure 19. Measured settling velocities for OSU SREC ASBR mixed liquor

Mixing Intensity Adjustment

During operating period 2, March 1, 2009 to July 17, 2009, the reactor operated at a 20 day HRT, one cycle per day, and an OLR of 0.37 kg VS/m^3/day (table 6). During this period, reactor temperature began to climb above the minimum 20˚C temperature due to environmental heating. The ASBR mixing system as designed provided 16.8 turnovers per day and a mixing intensity of 14 W m^{-3}. This mixing intensity is more than double the 6.6 W m^{-3} recommended by the USEPA (USEPA, 1979). Additionally, for ASBR's with mixed liquor solids concentrations of 5% or less, mixing at 8 W m^{-3}has no significant effect on methane yield (Karim et al, 2005). To reduce mixing intensity, the primary mixing pump (69 l/s) was taken offline, and only the heat exchanger mixed liquor recirculation pump was used for mixing. When heating was not required, slurry flowed through the heat exchanger without the boiler operating. This reduced the daily turnovers to 2 per day and reduced mixing intensity to 0.028 W m^{-3}. The ASBR was operated under these conditions until mid-July 2009 when incremental HRT reduction began.

HRT Reduction

The HRT was reduced from 20 days to the optimum of 5 days (Ndegwa et al., 2005, 2008) over a period of 4 months. The HRT was incrementally reduced at a rate of 1 day per week. When HRT reached 15 and 10 days, the ASBR was maintained at these HRT's for 4 weeks before further reductions continued. The incremental reduction of the HRT required changes to the influent volume, cycle length, and effluent standpipe selection. The effluent standpipes heights from the reactor floor are 3.20, 2.59, and 1.98 m (10.5, 8.5 and 6.5 ft.) (fig. 18). The lowest standpipe was not used due to concerns the unsupported membrane cover would be damaged at the lower liquid level. The daily manure production of 20.3 m^3 (5,400 gals) was sufficient for operation of the ASBR using the middle standpipe down to a HRT of 17 days. At an HRT of 16 days, the cycle length was reduced from 24 to 12 hours, giving two cycles per day. For HRTs of 15 days and less, both a cycle length of 12 hours and additional influent volume were required to maintain a settling distance of 0.3 m. The additional volume was created by pumping lagoon effluent into the lift station. In late November, 2009 the incremental HRT reduction operation reached the goal HRT of 5 days. This required addition of 48 m^3 of lagoon effluent to the influent manure stream, resulting in a total daily influent volume 68.3 m^3. Organic loading rate increased slightly with the addition of lagoon effluent as shown in table 6.

Steady 5 day HRT

Between December 7, 2009 and June 7, 2010, the ASBR was operated constantly at 5 day HRT, 2 cycles per day, and 0.50 kg VS / m^3 - day OLR. On June 7, 2010 the system was taken down for maintenance, ending the study period reported in this chapter.

Sampling and Analysis

Reactor influent, mixed liquor, and effluent samples were taken weekly during all operational periods. Influent manure samples were taken from the equalization tank via an access hatch at the top of the vessel using a Wheaton Science 0.76 l PVC coliwasa sampler. A port on the external heat exchanger shown in figure 19 provided the sampling location for the 1 l mixed liquor samples. Grab samples of effluent were taken from splitter box 2 (Figure 17) during effluent decanting using a 18 l bucket and 1 l samples were retained for analysis.

Sample total solids were measured by drying samples for 24 hours at 103˚C; volatile solids were measured by ashing the dried samples at 550˚C for 2 hours. Chemical Oxygen Demand (COD) was measured using CHEMetrics dichromate digestion vials and analyzed color metrically on a spectrophotometer. Sample pH was measured using an Accumet pH electrode (APHA, 1998).

Measurement of Digester Performance

Organic loading rate (OLR) in terms of COD or VS was calculated by dividing influent organic matter mass added over a given time period by reactor volume:

$$OLR = (C_{OI} * V_c * R)/V_{ro} \tag{14}$$

where:

OLR = Organic Loading Rate (kg OM m^{-3} day^{-1})

C_{OI} = Influent OM content (kg m^{-3})

V_c = Cycle volume (m^3)

R = Cycles per day (day^{-1})

V_{ro} = Reactor operating volume (m^3)

Hydraulic retention time was calculated using eq. 1. Solids Retention Time is determined by dividing the mass of the mixed liquor volatile solids by the mass of volatile solids leaving the reactor. This included both solids leaving in decanted effluent and wasted sludge:

$$SRT = (C_{OML} * V_{ro})/(C_{OE} * V_c * R + C_{OS} * V_{Sludge} / t_s) \tag{15}$$

where:

SRT = Solids retention time (days)

C_{OML} = Volatile solids concentration of mixed liquor (kg m^{-3})

C_{OE} = Volatile solids concentration of decanted effluent (kg m^{-3})

C_{OS} = Volatile solids concentration of wasted sludge (kg m^{-3})

V_{sludge} = Volume of wasted sludge (m^3)

t_s = sludge wasting period (day)

Sludge was not intentionally wasted during the study period; however, due to ASBR depth sensor malfunctions, solids wasting did occur as mixed liquor was released during under mixing during the react phase. These events were recorded by the ASBR automated control system, and based upon measured mixed liquor concentrations and reactor mixed liquor volume changes, the mass of solids wasted was determined. The running daily average of the unintentional wasted solids masses given in table 7 were tabulated and included in the calculation of the reactor's SRT as described in eq. 16.

Table 7. Volatile solids masses, volume, and dates of unintentional solids wasting events due to depth sensor malfunctions

Date	Volume	Volatile Solids Mass
	m^3	kg
2/4/09	20	63
2/5/09	20	64
3/17/09	20	89
6/29/09	20	84
7/20/09	40	233
1/11/10	34	142

$$SRT = (C_{OML}*V_{ro})/(\sum C_{OE}*V_c*R + \sum C_{OS}*V_{Sludge} / t_{SRT}) \qquad (16)$$

where:

V_{ML} = Volume of mixed liquor lost during depth sensor malfunction (m^3)

t_{SRT} = Time period for SRT calculation (days)

Reduction of organic matter in an ASBR digester occurs due to two processes: destruction through formation of biogas and removal by settling. When an ASBR is added to a liquid manure handling system, the wasted sludge will most likely be dried and marketed as concentrated nutrients, thus removing sludge organic matter from the liquid flow stream. The only organic matter added to downstream components is organic matter carried in decanted effluent. Therefore, the organic matter removal efficiency of an ASBR digester is measured by organic matter removed in the liquid stream divided by influent organic matter:

$$ORE = 100*(C_{OI}*V_C*R - C_{OE}* V_C*R)/(C_{OI}*V_C*R) \tag{17}$$

where:

ORE = Organic Matter Removal Efficiency (%)

The most relevant measure of organic matter conversion to methane in an ASBR digester is specific methane yield, or the volume of methane produced per mass of organic matter added. Specific methane yield was calculated in this study using eq. 18:

$$SMY = (Q_B*C_{MB}) / (C_{OI}*V_C*R) \tag{18}$$

where:

SMY = Specific Methane Yield (m^3 CH_4/kg OM)

Q_B = Volume of Biogas Produced (m^3/day)

C_{MB} = Biogas Methane Concentration (m^3 CH_4/m^3 biogas)

Another measure of digester performance is volumetric reactor efficiency:

$$VRE = (Q_B*C_{MB})/V_{ro} \tag{19}$$

where:

VRE = Volumetric reactor efficiency (m^3 CH_4 m^{-3} reactor day^{-1})

Although conversion efficiency in an ASBR digester may be quite high as measured by specific methane yield, volumetric methane production efficiency will be low compared to other reactors found in the literature due to the low strength influent treated in ASBR digesters. It is best to use VRE to compare an ASBR to other digesters treating similarly low strength wastes such as a covered lagoon digester. This is due to the low VOLR of the ASBR which results in VRE much lower than that observed in other digester systems such as a CSTR.

Results and Discussion

Influent characteristics are given in table 8. Masses of influent organic matter varied over the study period, but stayed within the range given in table 8. The pH of flushed manure was slightly more acidic than neutral. Concentration of organic matter and pH of ASBR influent decreased as lagoon effluent was added to the lift station in operational period 3c. When additional lagoon effluent was not added to the influent stream, influent pH ranged between 6.5 and 6.8. Volume of influent also increased as lagoon effluent was added, but OLR stayed relatively constant (table 6) due to the relatively small amount of organic matter contained in the lagoon effluent compared to flushed manure.

Mixed liquor (as sampled at the heat exchanger port) and decanted effluent quality characteristics are given for each operational period in tables 8 and 9. Both mixed liquor and effluent pH remained above 7.0 in each phase of operation regardless of changes to operation. Calculated masses of mixed liquor fixed, volatile, and total solids masses are given in figure 20. Dates of water level malfunction and loss of sludge are indicated by arrows in figure 20. Fixed solid mass remained relatively constant at 750 to 1,000 kg throughout the study period. This indicates a constant flow of inorganic material through the reactor with very little precipitation of salts. .

Table 8. Mixed Liquor Characteristics.

Operational Period	pH			TS			TSS			VS			VSS		
					$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$
	n	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**
1	11	6.9	0.03	11	5,900	611	6	3,900	199	11	3,900	275	6	3,500	250
2	28	7.0	0.1	32	6,600	1,100	20	4,100	220	32	4,400	360	20	3,500	250
3a	8	7.0	0.05	8	9,900	1,600	8	8,200	1,900	8	7,000	1,400	8	6,400	1,500
3b	6	7.0	0.1	6	9,800	1,700	6	8,100	2,200	6	6,800	1,600	6	6,300	1,800
3c	5	7.1	0.1	5	10,000	600	5	7,800	480	5	7,000	410	5	6,400	460
3d	5	6.9	0.1	5	10,000	940	5	8,200	1,000	5	7,300	950	5	6,800	970
4	5	7.0	0.1	5	6,900	930	5	5,100	1,000	5	4,700	780	5	3,900	300

Table 9. ASBR Decant Effluent Characteristics

Operational Period	pH			TS			TSS			VS			VSS		
					$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$		$mg\ l^{-1}$	$mg\ l^{-1}$
	n	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**	**n**	**X**	**SD**
1	8	7.0	0.04	8	4,800	430	8	2,900	645	8	2,900	375	8	2,300	420
2	27	7.1	0.1	27	4,500	935	15	2,800	600	27	2,600	760	15	2,400	350
3a	7	7.1	0.1	7	3,800	2,100	7	730	130	7	2,100	1,600	7	530	75
3b	6	7.1	0.1	6	3,500	2,000	6	640	120	6	1,800	1,400	6	460	50
3c	5	7.2	0.1	5	4,700	2,200	5	2,600	2,100	5	2,600	1,600	5	2,000	1,600
3d	5	7.2	0.1	5	3,700	1,100	5	1,800	920	5	1,900	870	5	1,400	770
4	5	7.1	0.1	5	3,300	1,300	5	2,200	1,300	5	1,700	880	5	1,700	1,200

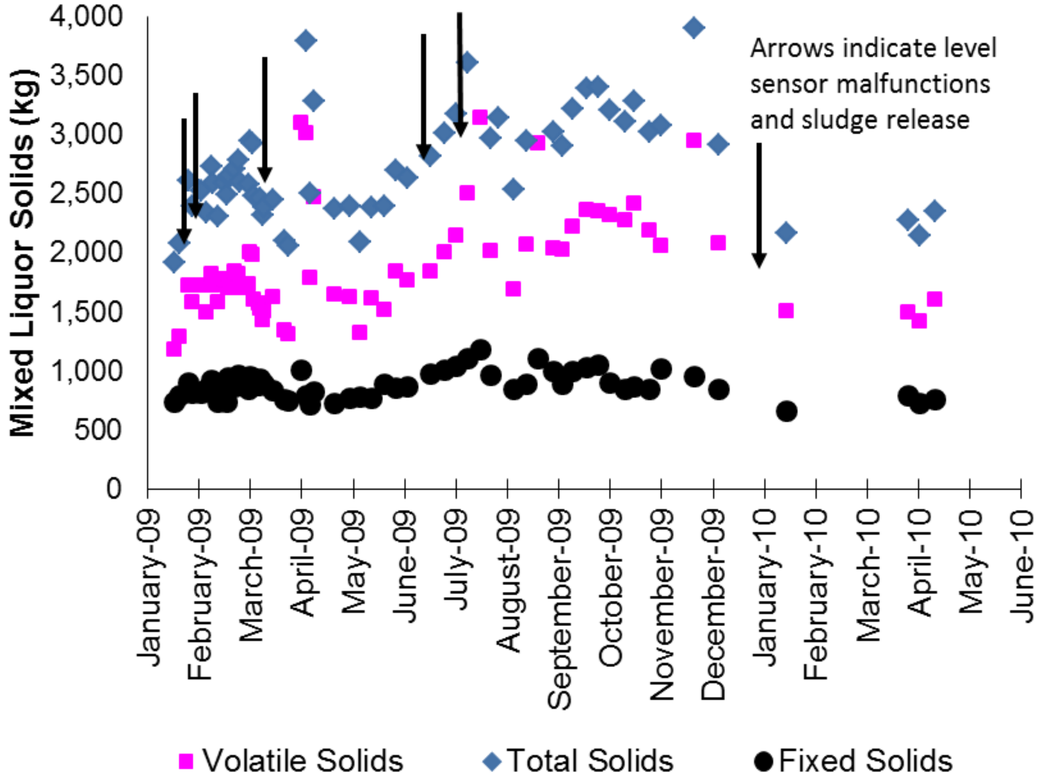

Figure 20. OSU SREC ASBR mixed liquor solids mass during continuous operation

Gas measurements were hampered by lack of a good metering system. The rotary displacement biogas meter failed to provide continuous or reliable measurements after August 2009. Meter failures were the result of the fine tolerances between the meter housing and rotor. Fine particulates in the biogas routinely fouled the meter. During freezing conditions moisture in the biogas resulted in ice in the meter during low gas production. Additionally, the rotary displacement meter requires the meter and gas piping to be independently supported, initial installation completed in such a manner causing the housing to torque and stall the rotor. The mounting supports for the meter piping were modified and provided some relief from meter failure however this did not provide a complete solution. A mass flow meter would be recommended to aid in more reliable gas metering as well as location of meter within an environmentally protected enclosure and biogas particulate filtration. The most reliable periods of gas production were during start-up and from April to September 09 (figure 21). No biogas production was observed from December 9, 2008 to February 2, 2009 while the ASBR was offline. Average biogas production rate during these periods was 80 m^3 day^{-1}.

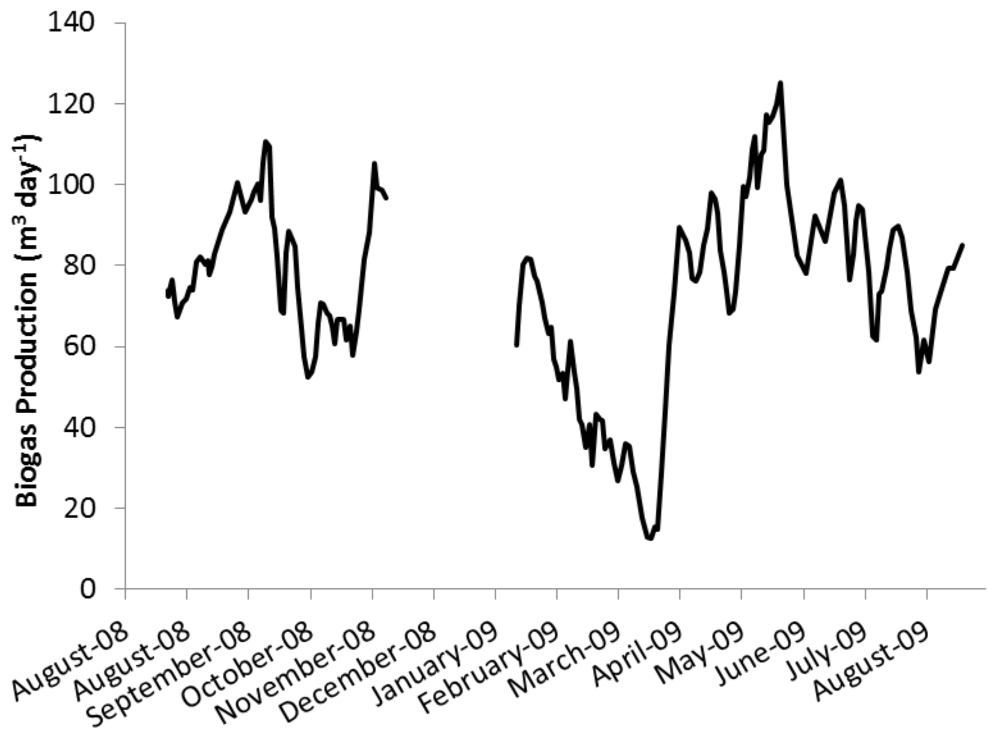

Figure 21. Daily biogas production during the first year of ASBR operation (5 day average)

Organic Matter Mass Balance

An organic matter balance in terms of VS was constructed on a daily time step for the continuous operation of the OSU SREC ASBR. Figure 22 shows the masses of COD flowing into and out of the digester. Solving the balance equation for the change in mixed liquor organic matter gives:

$$\Delta M_{MLO} = C_{OI} * V_c * C - C_{EO} * V_C * C - C_{SO} * V_{sludge} - UMY * C_{MB} Q_B \qquad (20)$$

where:

UMY = Ultimate Methane Yield (m^3 CH_4/kg VS destroyed)

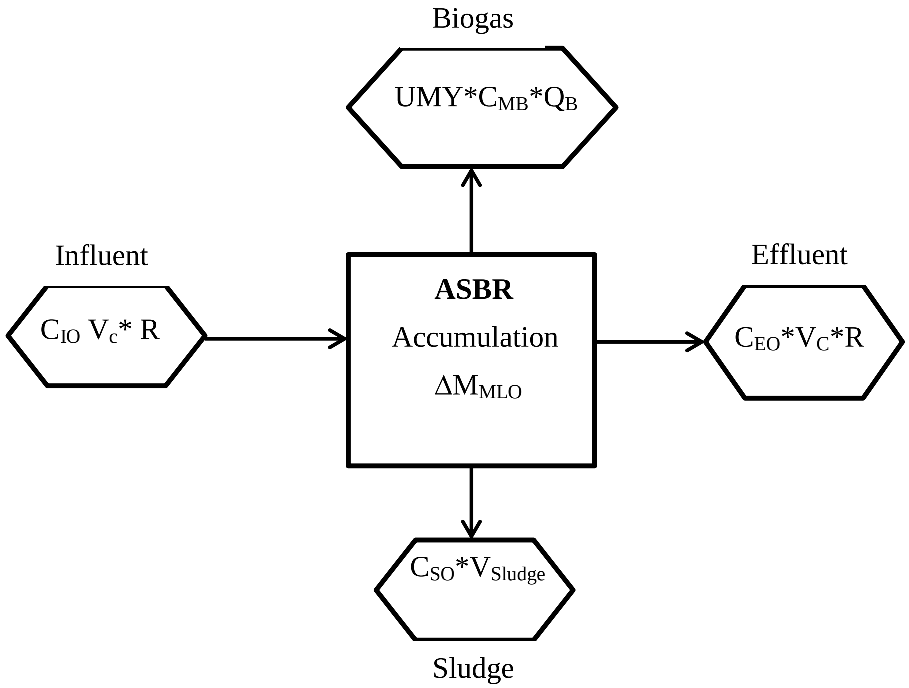

Figure 22. Mass balance of carbon across the reactor

During the time period April 25 to August 4, 2009, the only unknown in eq. 20 is the concentration of CH_4 (C_{MB}) in biogas. The mass balance was run on a daily time step during this period to calibrate for C_{MB} using measured values of both VS and COD. COD measurement were converted to VS using the average ratio of chemical oxygen demand to volatile solids; for swine manure the ratio was 1.8 g COD g^{-1} VS ($R^2 = 0.97$) and 1.6 g COD g^{-1} VS ($R^2 = 0.93$) for mixed liquor samples taken during the continuous operation. Ultimate methane yield is a constant ranging between 0.35 and 0.41 m^3 CH_4 kg^{-1} COD depending on reactor temperature and pressure (Smith, 1981). Using an ultimate methane yield of 0.35 CH_4 kg^{-1} COD and a COD to VS ratio of 1.8 g COD g^{-1} the UMY in terms of VS is 0.63 m^3 CH_4 kg VS^{-1}.

The mass balance was run using influent volume, influent VS, MLVS, effluent volume and effluent VS from April 25 to August 4, 2009. This gave an estimation of the biogas $C_{MB\ of}$ 0.65 m^3 CH_4 m^{-3} biogas. The linear correlation between the predicted and measure mixed liquor volatile solids mass for C_{MB} calibration period gave a R^2 value of 0.45 shown in figure 23. Using the estimated biogas methane concentration the mass balance was run for the entire operational period February 2, 2009 to June 6, 2010. The daily predicted mixed liquor VS masses are shown with the corresponding measured values in figure 24. A linear correlation between measured and predicted mixed liquor VS masses gave an R^2 value of 0.60. Predicted daily biogas production versus measured values is given in figure 25, the linear correlation between predicted and measured daily biogas production values gave an R^2 value of 0.44 shown in figure 26.

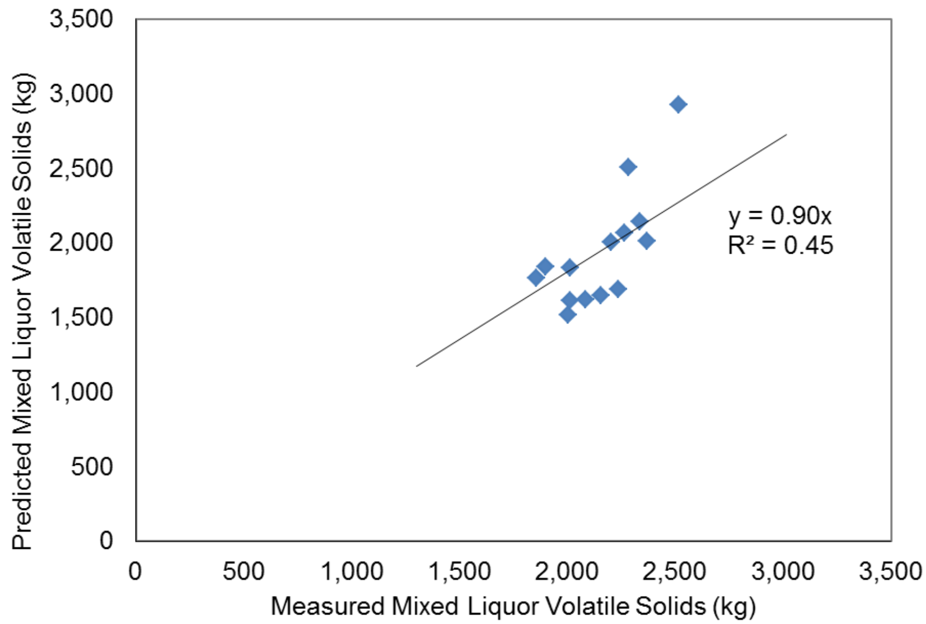

Figure 23. Linear correlation for measured and predicted mixed liquor volatile solids mass for organic matter mass balance calibration period

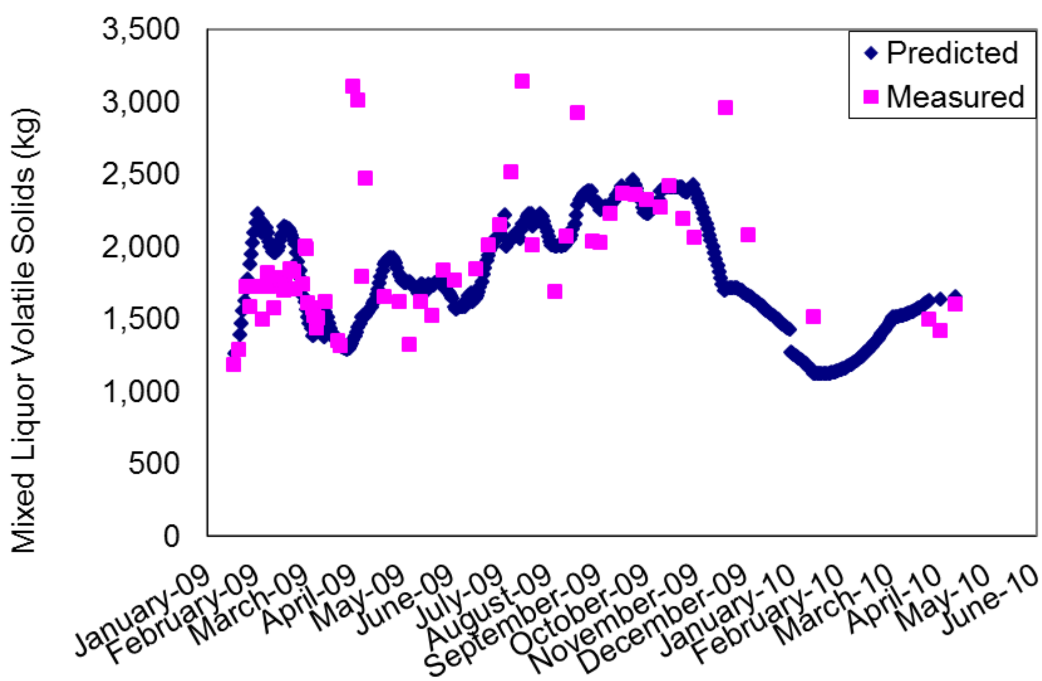

Figure 24. Comparison of measured and predicted mixed liquor volatile solids for OSU SREC ASBR

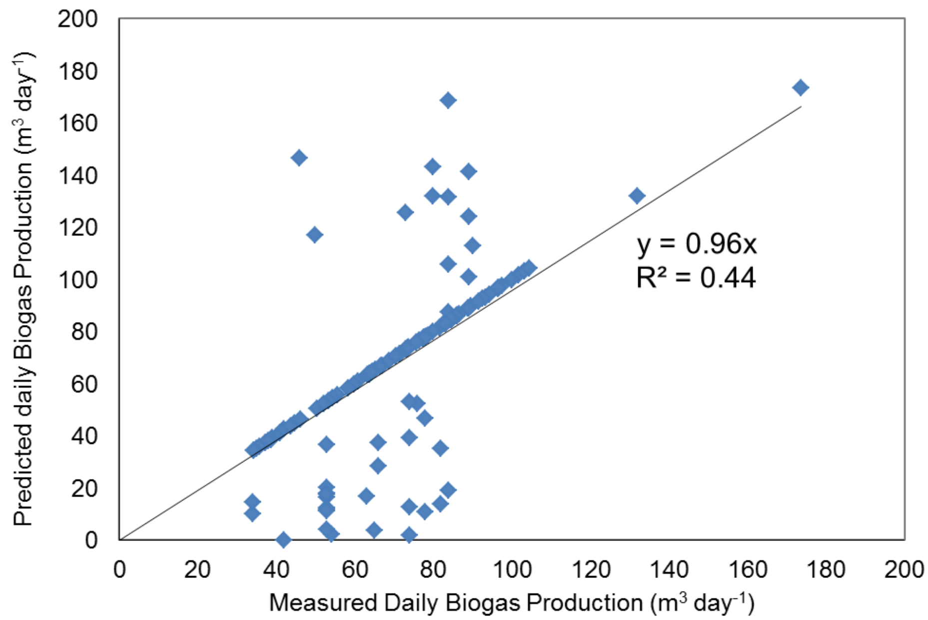

Figure 25. Linear correlation for measured and predicted daily biogas production for organic matter mass balance

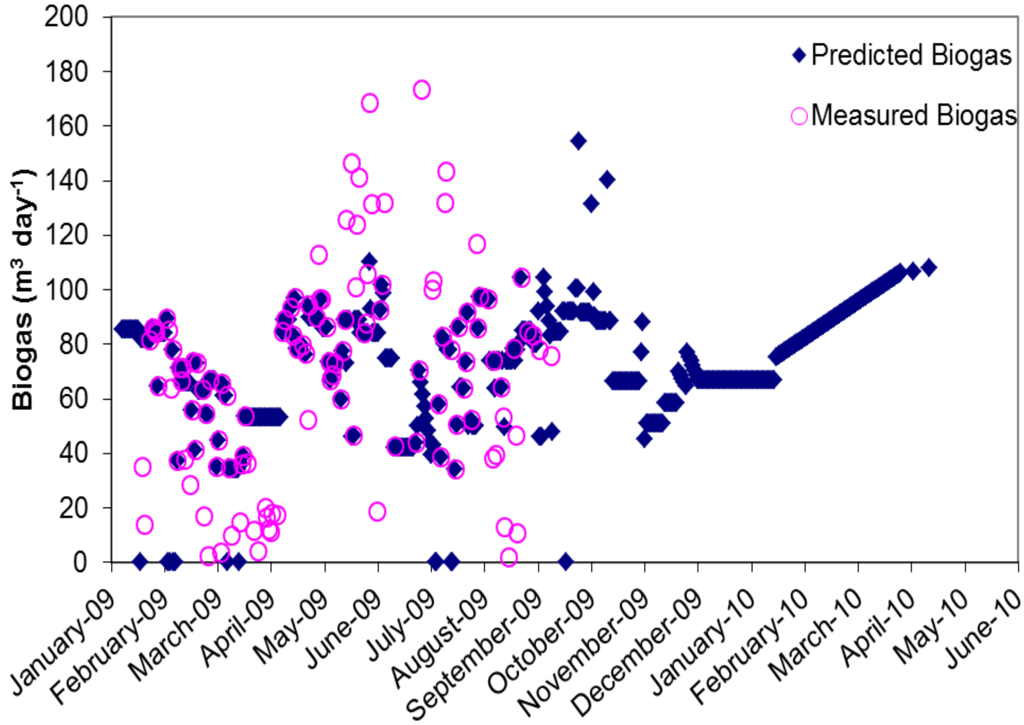

Figure 26. Comparison of measured and predicted biogas production for OSU SREC ASBR

Average SRT, organic matter removal efficiency (VS and COD) specific methane yield, and volumetric reactor efficiency for each operational period was calculated (table 10). The mass balance was used to provide biogas production values. Based upon the data provided in table 10 operational results of the full scale reactor were comparable to the lab scale ASBR's operated by Ndegwa et al., 2003 and Ndegwa et al., 2005. Based upon the mass balance analysis, the reactor biogas methane content was estimated to be 65% throughout operation; this is within 10% of the biogas composition range observed in both lab scale studies of 65 to 70%. Although the biogas methane composition was slightly lower than the lab scale reactors, the Volumetric Reactor Efficiency for the 4 d

HRT single and multiple cycles per day experiment, 0.12 l CH_4 l^{-1} day l^{-1}, was similar to those estimated for the full scale reactor 0.11- 0.16 l CH_4 l^{-1}-day l^{-1} (Ndegwa et al., 2005).

The SMY's for periods 1 and 3b are lower and higher, respectively, compared to the other operational periods (table 10). The SMY for period 1 is 65% of that observed for period 3b. Using the modified Chen and Hashimoto equations (eqs. 10, 11, and 12) the specific methane yield was estimated for the periods using the influent volatile solids concentration, HRT, SRT and VRE given in tables 4, 6, and 10. Reactor temperatures for these two periods were 24 and 30.5°C, respectively. Solving the modified Chen and Hashimoto equations for SMY (B_o), SMY values for the periods 1 and 3b were calculated 0.31and 0.48 m^3 CH4 kg VS^{-1}. Similar to observed SMY's the SMY of period 1 was 64% of SMY for period 3b, indicating that observed difference in SMY for the two periods is agreement with Contios model kinetics.

Table 10. Digester performance measures based on measured values and organic matter mass balance.

Operational Period	HRT	SRT	Organic Matter Removal Efficiency		Biogas Production	Specific Methane Yield	Volumetric Reactor Efficiency
			VS	COD			
	days	days	%	%	m^3 day^{-1}	m^3 CH_4 kg VS^{-1}	m^3 CH_4 m^{-3} day^{-1}
1	20	33	66	70	69**	0.26**	0.12**
2	20	47	60	65	65**	0.35**	0.11**
3a	20-17	35	74	73	71	0.36	0.12
3b	16-15	37	64	62	76	0.40	0.15
3c	15-10	43	64	65	78	0.35	0.16
3d	10-5	35	63	63	69**	0.33**	0.14**
4	5	44	64	59*	81**	0.33**	0.15**

*Estimated values based on sampled VS and COD to VS ratio of 1.8 for swine manure and 1.6 for mixed liquor

**Estimated values based upon the organic matter mass balance results

Effluent suspended solids concentrations of the full scale reactor were up to three times higher than those reported in Ndegwa et al, (2008). However the increase in effluent suspended solids would be expected as the influent solids of the full scale reactor was significantly higher than that of the lab scale, with average TS concentrations ranging from 5,900 to 10,000 mg/l compared to 3,560 mg/l (Table 8). Similarly, the COD reductions for the full scale reactor were lower than that observed for the lab scale reactors, 65 - 73%, compared to 80 to 85%.

Based upon biogas production, this difference and the full scale consistent ability to maintain an SRT of 30 days or greater indicates that this difference is due to settling rather than microbial degradation of the influent. The difference in settling effectiveness between the full scale and lab scale results are partially due to the dynamic influent solids concentration of the full scale reactor. Although both experiments utilized swine manure from the same production facility, Ndegwa et al. (2005) collected and adjusted manure samples prior to digestion to maintain a consistent influent manure solids concentration. Although the dynamic influent solids concentration does appear to have increased the effluent solids concentration of the full scale reactor, the SRT was maintained above 30 days. The maintaining of a SRT of greater than 30 days provided the full scale reactor the ability to maintain a relatively stable organic removal and specific methane yield throughout operation. The importance of maintaining a high SRT is shown in the full scale experiment through stable performance regardless of reactor operation parameter changes to temperature during summer months, mixing intensity reduction, and HRT reduction.

Conclusions

The cold start and continuous operation of the full scale OSU SREC ASBR shows that anaerobic digestion of low strength swine manure is achieved using the ASBR. The ability to utilize the cold start method provides increased flexibility in the startup process by reducing the need for seed sludge volumes. However, the overall start up time is increased compared to a seeded reactor as additional time is required to achieve a mixed liquor solids content that promotes settling and provides adequate microbial biomass. The results of this study also show that the ASBRs treating low strength swine manure can be started without the aid of solids inoculum and operated at an HRT of 5 days after an adequate start period.

The OSU SREC reactor operated as a full scale reactor as part of a functional swine operation and is capable of achieving performance similar to that of lab scale models. Biogas and methane production of the full scale reactor, based upon volumetric production rates, were found to be scalable for this manure source and reactor operational parameters, 0.12 m^3 CH_4 m^{-1} day^{-1}. Settling is the critical physical parameter to ASBR performance. With similar biogas production rates, the difference in effluent organic removals between lab and full scale reactors appears to be a result of differences in settling ability of reactors. Although settling may have resulted in lower organic removal rates (65 to 70 % compared to 80 to 85% in the lab scale reactors) the SRT was consistently maintained above 30 days. The reduction of HRT's from 20 to 5 days did not reduce the ASBR's ability to treat low strength swine manure in a full scale application. The similar results of the full and lab scale reactors indicate the potential for the ASBR to be included as part of a comprehensive waste management system for swine producers.

CHAPTER IV

CO-DIGESTION OF CRUDE GLYCEROL IN AN ANEROBIC SEQUENCING BATCH REACTOR (ASBR) FED LOW STRENGTH SWINE MANURE

Introduction

Anaerobic digestion as a waste treatment process allows for nutrient and energy recovery from aqueous and water soluble organic waste streams. The high level of available nutrients and organic matter in livestock manures makes it an ideal candidate for use in anaerobic digestion. In addition to providing adequate nutrients for biomass growth, manure also provides needed alkalinity for process stability. For typical wet anaerobic digestion of animal manures, solids contents of less than 40% can be used in continuous stirred and plug flow type reactors (Ward et al., 2008). However, at high solids contents, settling of suspended matter, specifically biomass, is not applicable. Thus effluent solids concentrations are equal to mixed liquor solids concentration, requiring hydraulic retentions times of 20 to 30 days to achieve equivalent solids retention times.

Swine manure, unlike beef and dairy manure, is quite dilute, less than 10% solids, which doesn't not allow for digestion in plug flow digesters. In CSTR reactors, adequate mixing is required to maintain suspension of solids, as settling more readily occurs at lower solids concentrations. Due to the dilute nature of swine manure, it's an ideal feedstock for use in an Anaerobic Sequencing Batch Reactor (ASBR).

The use of an ASBR for the waste treatment of dilute swine manure has been shown effective despite the dilute nature of the swine manure (Steele and Hamilton, 2010). The low strength of the swine manure compared to other manure sources provides an opportunity for co-digestion in an ASBR while still maintaining a low solids feedstock concentration. One co-digestion feedstock in particular that is well suited for anaerobic digestion in an ASBR is waste glycerol from the production of biodiesel. Crude glycerol is produced at a rate of 10% per weight of biodiesel. Crude glycerol must be removed from the biodiesel due to emission of toxic gases from burning.

The energy density of crude glycerol makes it a valuable energy feedstock. With an energy density of 120,000 Btu/gallon it is comparable to many fossil fuels as a fuel oil. As an energy fuel source there are significant drawbacks to its use: it has a high viscosity, high ignition temperature, and produces a toxic combustion byproduct – acrolien gas. These problems have largely stymied the use of glycerol as an inexpensive energy source (R. Scott Frazier, OCES Renewable Energy – Energy Conservation Specialist. Personal Communication, February 10, 2010).

With a chemical oxygen demand of 800 to 1,400 g l^{-1}, a relatively small volume of glycerol is required for significant biogas production increases. Biodegradable, water miscible, and containing few particulate solids, minimal changes are required for use of crude glycerol as a co-digestion feedstock. Anaerobic digestion laboratory trials with reaction periods of less than 40 hours produced complete conversion of glycerol to methane (Lopez et al., 2009, Ma et. al., 2008). Wohlgemut et al. (2011) and Fountoulakis et al. (2010) found that 1% by volume additions of glycerol to continuous stirred reactors (CSTR) operating at 20 day HRTs doubled biogas production. However, from the literature available, 1% by volume addition of glycerol has been the maximum addition for sustained stable operation. The results of these studies do not indicate an exact reason for the 1% maximum addition.

Objectives

- Determination of the maximum inclusion rate of crude glycerol from biodiesel production for stable co-digestion in an ASBR treating low strength swine manure at a 5 day HRT and 20°C operating temperature.
- Determination of the digestibility of crude glycerol and methane yield for co-digestion at the maximum inclusion rate for stable co-digestion in an ASBR treating low strength swine manure at a 5 day HRT and 20°C operating temperature.

Methods and Materials

Materials Used

Fifty-five liters of crude glycerol were donated by Murray Thibodeaux. Mr. Thibodeaux is a small-scale producer of biodiesel. His main source of raw materials is waste grease from fast food restaurants in Tulsa, Oklahoma. Characteristics of the crude glycerol utilized throughout this study are given in table 11, sampled and measured August 17, 2010. Swine manure was collected on a weekly basis from the Oklahoma

State University Swine Research and Education Center (OSU SREC). Characteristics of manure are also given in table 11.

Table 11. Characteristics of Crude Glycerol and Swine Manure.

	Crude Glycerol	Swine Manure (n=23)	
		X	SD
TS g l^{-1}	-	8.3	4.9
VS g l^{-1}	580	6.0	4.0
COD g l^{-1}	1,400	12.3	5.8
pH	9.6	7.5	0.5
Total N mg l^{-1}	340	920	260
Total P (as P_2O_5) mg l^{-1}	79	545	330
Total K (as K_2O) mg l^{-1}	210	440	110
Ca mg l^{-1}	4.5	360	235
Mg mg l^{-1}	2.5	120	60
Na mg l^{-1}	8,200	130	29
S mg l^{-1}	13	88	45
Fe mg l^{-1}	26	30	27
Zn mg l^{-1}	0.6	19	14
Cu mg l^{-1}	2.7	2.5	1.8
Mn mg l^{-1}	0.4	5.6	4.0

As can be seen in table 11, the composition of the swine manure collected from OSU SREC was highly variable. Standard deviations are nearly on the same scale as the averages. This level of variability is to be expected on a working farm. Also note the swine manure is fairly low in solids – 8.3 g TS/L, which roughly translates to 0.83% TS by weight. The crude glycerol is much higher in energy content than dilute swine manure. Chemical oxygen demand is the chief measure of energy in anaerobic systems. Crude glycerol’s COD is 116 times that of the swine manure. The plant nutrient content (N, P, K) of crude glycerol is lower than swine manure.

Model Reactors

Model ASBR reactors with an operating volume of 18 l operated at 20˚C with a 5 day HRT were constructed of 6.5 gallon food grade buckets. Each reactor contained an internal coil for circulation of water for temperature control and a single downward jet mixing nozzle providing a mixing intensity of 0.028 W/m^3. The design and setup of the model ASBR's replicate the conditions found in the farm scale ASBR located and operated at the Oklahoma State University Swine Research and Education Center (SREC). Swine manure was collected from the OSU SREC and stored in a 200 l storage tank which was maintained at 10˚C and mixed prior and during feeding. An additional 14.4 l storage vessel was constructed for the glycerol treated manure which replicated the manure equalization and storage tank at the OSU SREC. Daily manure and glycerol volume were added to the glycerol treatment storage vessel as required to maintain a consistent volume of 14.4 l.

Test Procedures

Reactors were seeded using mixed liquor from the OSU SREC treating dilute swine manure. Swine manure was fed to the reactors at 3.6 l per day providing an average organic loading rate (OLR) of 2.4 g COD /l-day until stable biogas production was observed. The control reactor was maintained at these operating conditions for the duration of the study. The glycerol treatment reactor was fed 3.6 l per day of a manure-glycerol mixture from the 14.4 l manure storage tank. The glycerol content of the manure-glycerol mixture was increased step-wise until biogas production or volatile fatty acid (VFA) concentration of the treatment reactor increased above that of the control. Upon the observation of an upset condition, glycerol concentration was reduced step-wise one step until recovery of biogas or VFA concentrations. Upon recovery, glycerol inclusion increased at a rate of one half of the previous rate. Utilizing a step-wise incremental glycerol inclusion rate, the maximum inclusion rate of crude glycerol in the feed was determined.

Analytical Methods

Feedstock and effluent parameters were monitored bi-weekly and daily biogas volume measurements were made. Biogas production gas meter-logger added biogas flow every 15 minutes. Total Solids (TS) analysis was performed by drying samples at 103˚C for 24 hours and volatile solids (VS) were measured by drying samples at 103˚C for 24 hours followed by ashing dried samples at 550˚C for 2 hours. An Accumet pH electrode was used for sample pH measurement and titrations. Chemical Oxygen demand (COD) measurements were conducted using CHEMtrics dichromate digestion vials and analyzed colormetrically with a spectrophotometer (APHA, 1998). A two point VFA titration using the method described in Anderson and Yang, 1992, was used for VFA monitoring during the second glycerol volume rate increase. Samples for nutrient analysis of influent, effluent, and crude glycerol were analyzed by the Oklahoma State University Soil, Water, and Forage Analytical Laboratory.

Results and Discussion

Organic loading rate and daily biogas production for both control and treatment reactors are given in figures 27 and 28. Day 0 indicates the time at which glycerol addition began on the treatment reactor. The experiment lasted 257 days. The organic loading rate (solid line) shown in both figures is the manure+glycerol rate. As can be seen in these figures, the greatest variability in loading rate came from variability of the manure.

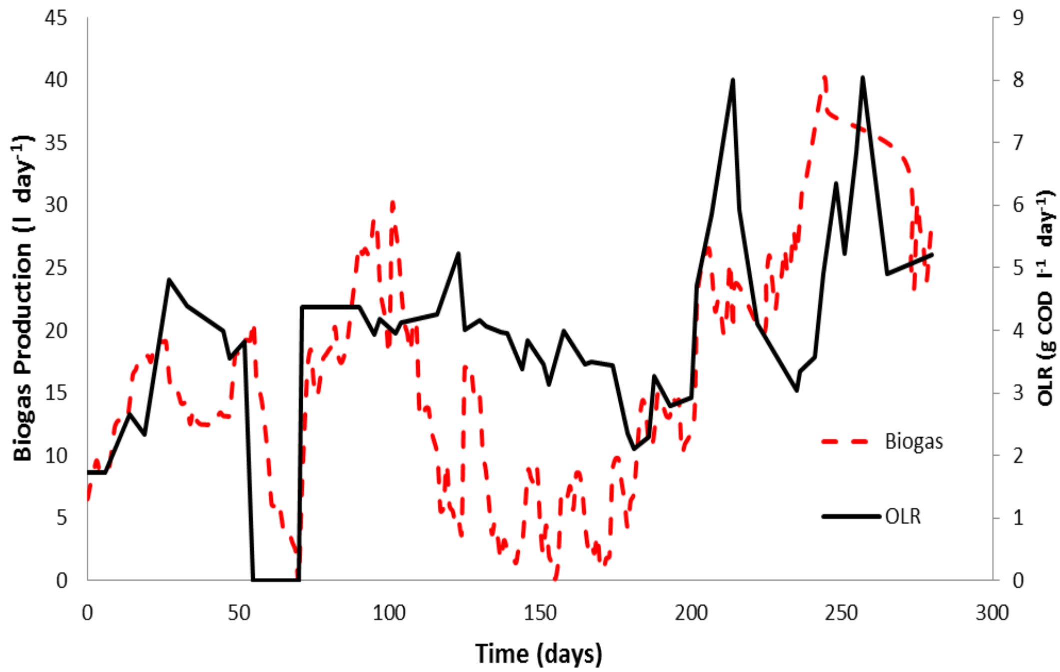

Figure 27. Organic loading rate (manure + glycerol) and biogas production in treatment reactor

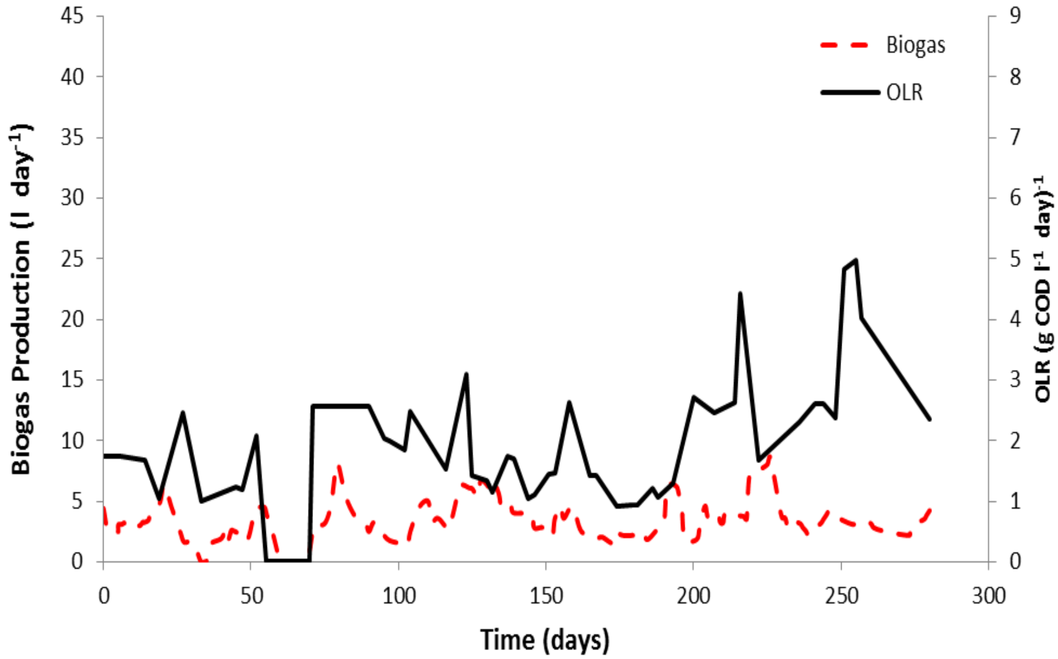

Figure 28. Organic loading rate (manure) and biogas production in control reactor

Daily biogas production versus percent by volume glycerol inclusion (% v/v) is shown in figure 29. The biogas production response to the step-wise glycerol inclusion is illustrated. During the first 50 days a rapid increase in biogas production, peaking at approximately 15 liters per day, was observed (fig. 27). At a glycerol inclusion rate of 1.25% biogas production dropped and recovered when dropped to 1.1% but crashed upon being raised to 1.2%. Feeding of both the treatment and control reactor was stopped for 15 days, then restarted at a glycerol inclusion rate of 1.1%. Biogas production returned and increased. The inclusion rate was increased form 1.1% to 1.22% and biogas production declined. The inclusion rate was reduced to 1.0% and the biogas production continued to decline. Review of the VFA concentration showed that the concentration had peaked at 2,050 mg/l as Acetic Acid (HAC) on day 121, nearly 10 times that of the control reactor (figure 29).

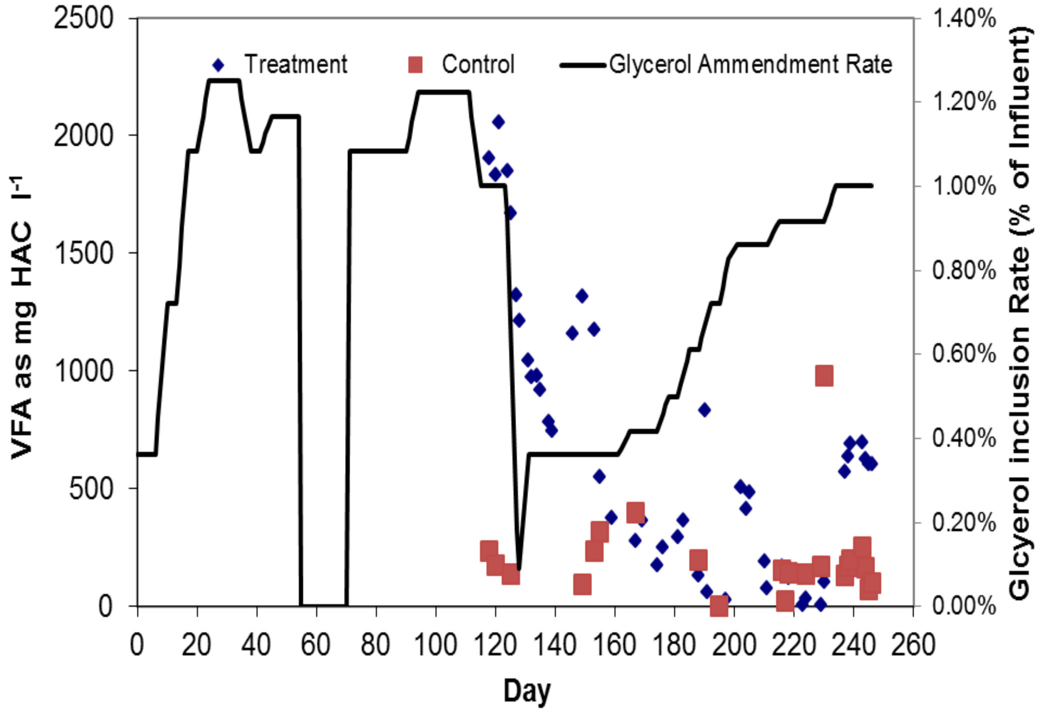

Figure 29. Control and glycerol treatment reactor VFA concentration and corresponding glycerol inclusion rate.

Glycerol inclusion was withheld for 3 days (days 125-127), and then resumed at 0.36% on day 127 for 34 days. On day 161 the glycerol inclusion rate was increased to 0.42%. Treatment reactor VFA had fallen to below 400 mg l^{-1} as HAC and remained between 200 and 400 mg l^{-1} as HAC, for two weeks. With VFA maintained below 500 mg/l as HAC stepwise glycerol inclusion was resumed increasing the inclusion rate approximately 0.1% each week until reaching 1.0% on day 231. The peaks in VFA concentrations at day 193 and around day 240 correspond to spikes in OLR and were not considered upset conditions.

It appears that the maximum inclusion rate of crude glycerol from biodiesel production for ASBR digesters is 1.1% of influent volume. The inclusion rate for stable, continuous operation for an ASBR is 1.0%. These results correspond with earlier findings for CSTR's (Wohlgemut et. al., 2011). The average biogas production at 1.0% inclusion was 29 l day^{-1}, which compares to a biogas production rate of 4.4 l day^{-1} for the control reactor. Table 12 shows the average biogas quality data for the treatment and control reactors. At a 1.0% glycerol inclusion rate, methane content was 73%. Methane content of biogas produced by the control reactor was 66%. Methane production increased by a factor of 7.3 at an 1.0% glycerol inclusion rate. Volumetric CH_4 production efficiency was 0.16 l CH_4 l reactor^{-1} day^{-1} for the control reactor and 1.2 l CH_4 l reactor^{-1} day^{-1} at the 1.0% glycerol inclusion rate.

Table 12. Biogas Quality in Control and Treatment Reactors.

	Control (n=4) (% by volume)		Treatment (n=4) (% by Volume)	
	X	SD	X	SD
CH_4	66	1.2	73	1.7
CO_2	31	1.3	23	2.0
H_2	0	0	0	0
N_2	3	0.4	4	0.2

Table 13 gives specific methane yield and % COD converted to CH_4 for the crude glycerol and swine manure used in this experiment. Methane produced by glycerol alone in the treatment reactor was 18.5 l CH_4 day^{-1} (21.5 l CH_4 day^{-1} in treatment reactor minus 3.0 CH_4 L day^{-1} in the control). Since the organic matter loading from glycerol was 50.4 g COD day^{-1} specific methane yield of glycerol was, therefore, 0.37 L CH_4 g^{-1} COD. Given that the ultimate methane yield, or the maximum theoretical volume of methane produced per g COD removed is 0.40 L CH_4 g^{-1} COD (Smith, 1981), then 92.5% of COD contained in glycerol was converted to methane by the ASBR digester. Specific methane yield per mass of volatile solids (0.89 L CH_4 g^{-1} VS) is similar to results found for food grease in biochemical methane potential assays (Moody et al., 2011).

Table 13. Organic Matter Conversion Factors for Swine Manure and Crude Glycerol determined from Control and Treatment Reactors.

	Manure	Crude Glycerol
Specific Methane Yield (L CH_4 g^{-1} COD)	0.069	0.37
Specific Methane Yield (L CH_4 g^{-1} VS)	0.14	0.89
COD Converted to Methane (%)	22.5	92.5

Organic matter removal efficiency of the treatment and control reactor throughout the 257 day testing period is given on the basis of COD in figure 30. Figure 31 shows the organic matter removal efficiency on a VS basis. Organic removal efficiency of the treatment reactor was higher and more consistent as the 0.92% inclusion rate was reached after day 214 (figure 30). COD removal was approximately 80% and VS removal was between 60 and 70. The removal rates are higher in the treatment reactor due the higher digestibility of crude glycerol. Even at higher organic removal rates, the effluent concentrations will be higher than an ASBR fed swine manure only, because the OLR is two to three times higher in the treatment rector. The increased average VS concentrations for glycerol treatment, 4,900 mg VS l^{-1} compared to the control, 1,400 mg VS l^{-1}, is shown in table 14.

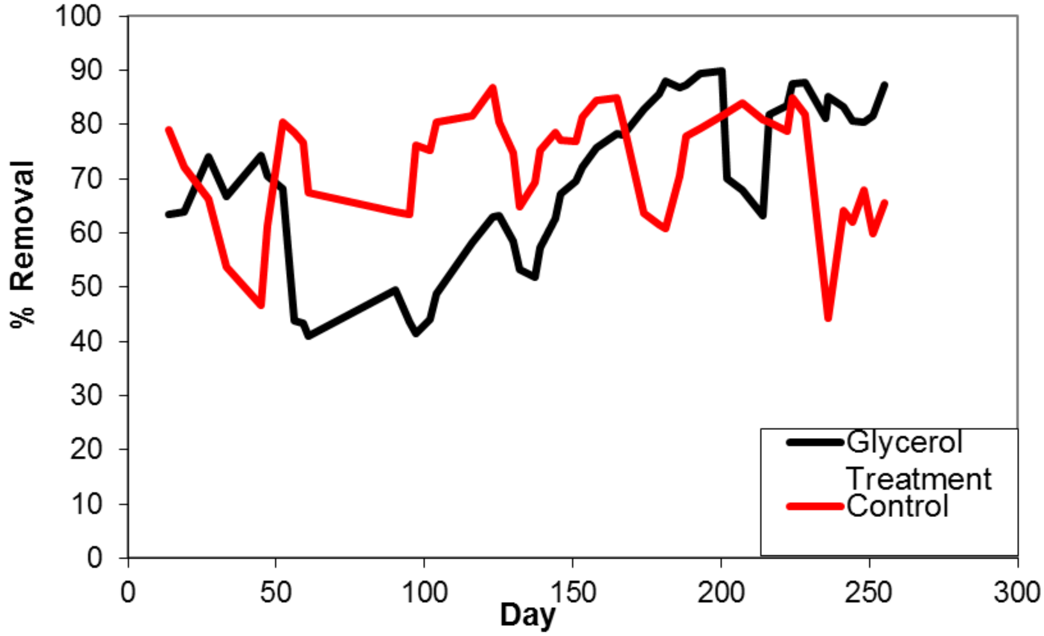

Figure 30. COD Removal Efficiency of Control and Treatment Reactors (3 point average).

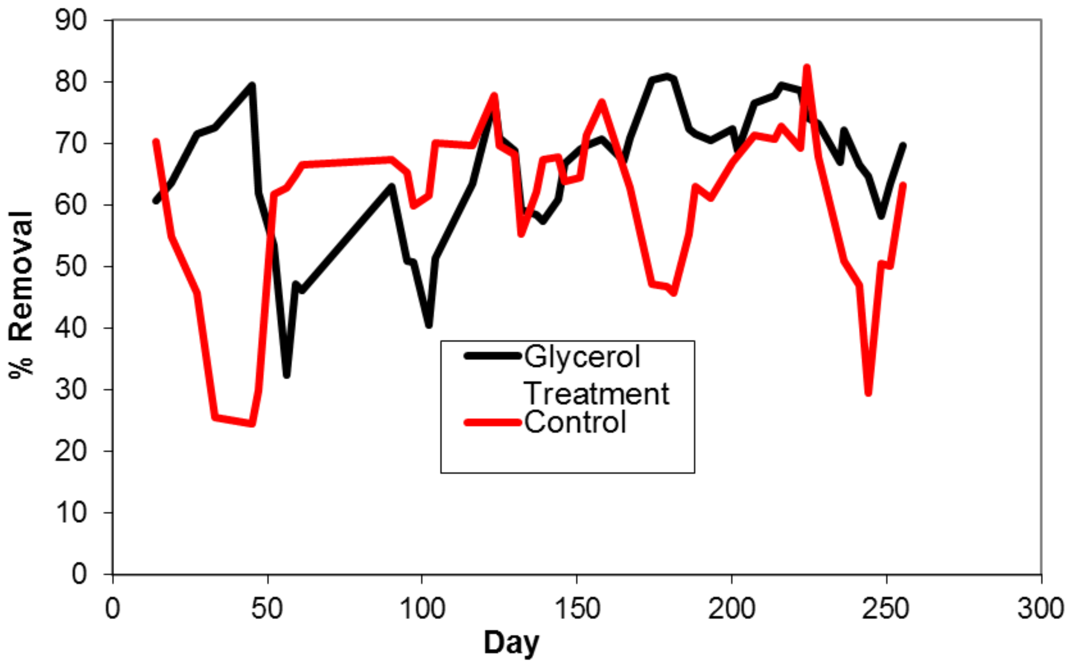

Figure 31. VS Removal Efficiencies of Control and Treatment Reactors (3 point average)

Sludge and Effluent Analysis

During the steady state period utilized for mass balance calculations, two sludge wasting events were employed; 2/21/12 and 3/27/12. During the sludge wasting, the mixing was continued during the settling phase, thus allowing for an unsettled cycle withdrawal to occur. The implementation of sludge wasting is an integral part of the steady state operation of the ASBR, by providing a mechanism for maintaining a consistent mixed liquor solids concentration. Tables 14 and 15 gives the solids distribution between the reactor's effluent and the sludge recovered during this period. The sludge solids concentrations were calculate by subtracting the average effluent solids concentration for the period. The solids mass was determined by multiplying the average concentration by the daily effluent volume (3.6 l) by the total number of cycles during the period (57). The sludge mass was determined based upon a sludge wasting volume of 3.6 l for each of two wasting events. For the glycerol treated reactor, 19 and 22% of total solids and total volatile solids were recovered as sludge, compared to 3 and 4% for the control reactor.

Table 14. Model ASBR effluent during steady state period

	Glycerol Treatment		Control	
	Effluent mg l^{-1}		Effluent mg l^{-}	
Date	Total Solids	Total Volatile Solids	Total Solids	Total Volatile Solids
1/19/2012	5,500	3,800	2,900	1,500
1/31/2012	5,400	3,700	2,900	1,700
2/28/2012	7,000	5,000	2,300	1,000
3/6/2012	6,300	4,300		
3/13/2012	10,300	7,600	3,900	2,400

Table 15. ASBR effluent and sludge solids distribution for steady state period

	Glycerol Treatment		Control	
	Effluent mg l^{-1}		Effluent mg l^{-1}	
	Total Solids	Total Volatile Solids	Total Solids	Total Volatile Solids
Effluent Average (mg/l)	6,900	4,900	2,700	1,400
Sludge Cycle Discharge-2/21/12 (mg/l)	33,800	27,500	5,700	3,600
Sludge - 2/21/12 (mg/l)	26,900	22,600	3,000	2,200
Sludge Cycle Discharge - 3/27/12 (mg/l)	44,000	35,800	3,900	2,400
Sludge - 3/27/12 (mg/l)	37,000	30,900	1,200	980
Total Cycles	57	57	57	57
Total Effluent Discharge (g)	1,400	1,000	550	290
Total Sludge Recovered (g)	230	190	15	11
Percent Recovered as Sludge (%)	16%	19%	3%	4%

Although there is an increased amount of solids produced and discharged from the glycerol treatment reactor, there is no increase in the mass of nutrients available in the sludge and effluent (table 16). This is a result of equivalent influent nutrient masses for both reactors. The change in sludge and solids production is a result of the increased organic loading of the glycerol treatment reactor, which allows for increased metabolic activity and growth compared to the control reactor. Based upon the values in table 14 it is estimated that volatile solids production for the glycerol treatment reactor is four times that of the control reactor based upon total sludge recovered.

Table 16 -ASBR sludge and effluent nutrient distribution

	Glycerol Treatment			Control		
	Total N	Total P	Soluble P	Total N	Total P	Soluble P
	g (avg.)	g (avg.)	g (avg.)	g (avg.)	g (avg.)	g (avg.)
			Period 1			
Effluent	45 (2.2)	30 (1.4)	2.4 (0.12)	56 (2.7)	10.8 (0.51)	2.8 (0.13)
Wasting Discharge	8.2	6.4	0.45	15	17.91	0.54
Sludge	6.2	4.9	0.33	12	17.40	0.41
Percent Nutrient Recovered in Sludge	**12%**	**14%**	**11%**	**17%**	**60.61%**	**12.3%**
			Period 2			
Effluent	93 2.7)	35 (1.0)	5.0 (0.15)	77 (2.3)	19.1 (0.58)	4.1 (0.13)
Wasting Discharge	8.3	8.0	0.62	2.8	10.02	0.13
Sludge	5.6	6.9	0.47	0.44	9.44	0.0
Percent Nutrient Recovered in Sludge	**5.5%**	**16%**	**8.5%**	**0.55%**	**32.46%**	**0.0%**
			Total			
Discharged	155	79	8.5	150	58	7.6
Sludge	12	13	0.8	12	27	0.41
Percent Nutrient Recovered in Sludge	**7.6%**	**16%**	**9.4%**	**8.2%**	**46.4%**	**8.4%**

The theoretical biomass production for the anaerobic digestion of glycerol is 0.041 moles of biomass ($C_5H_7NO_2$) per mole of glycerol digested. In terms of COD, biomass production from the digestion of has a theoretical yield of 0.041 g of biomass per g COD as given below.

$$1\ C_3H_8O_3 + 0.663\ NH_3 \rightarrow 1.648\ CH_4 + 0.526\ CO_2 + 0.041\ C_5H_7NO_2 + 0.622\ NH_4HCO_3$$

(113 g $C_5H_7NO_2$ / mol $C_5H_7NO_2$) x (0.041 mol $C_5H_7NO_2$ / 1 mol $C_3H_8O_3$) = 4.63g $C_5H_7NO_2$ / mol $C_3H_8O_3$

For the model ASBR, the daily 1% glycerol inclusion with a COD load of 49.86 g COD /day yields increases the reactor's theoretical biomass production by 2.04 g per day compared to the control as calculated below.

(4.63 g $C_5H_7NO_2$ / mol $C_3H_8O_3$) x (1 mol $C_3H_8O_3$ / 112 g COD) = 0.041 g $C_5H_7NO_2$ / g COD glycerol

(49.86 g COD glycerol / day) x (0.041 g $C_5H_7NO_2$ / g COD glycerol) = 2.04 g biomass / day

Mass Balances

A mass balance of the volatile solids was completed for the steady state period of glycerol inclusion utilized for sludge removal and nutrient analysis. For the glycerol treatment reactor, 89% of the influent volatile solids are accounted for in sludge, effluent and gas production, as shown in figure 32. For the control reactor, only 85% of the volatile solids were accounted for in the sludge, effluent, and biogas during the same period (figure 33).

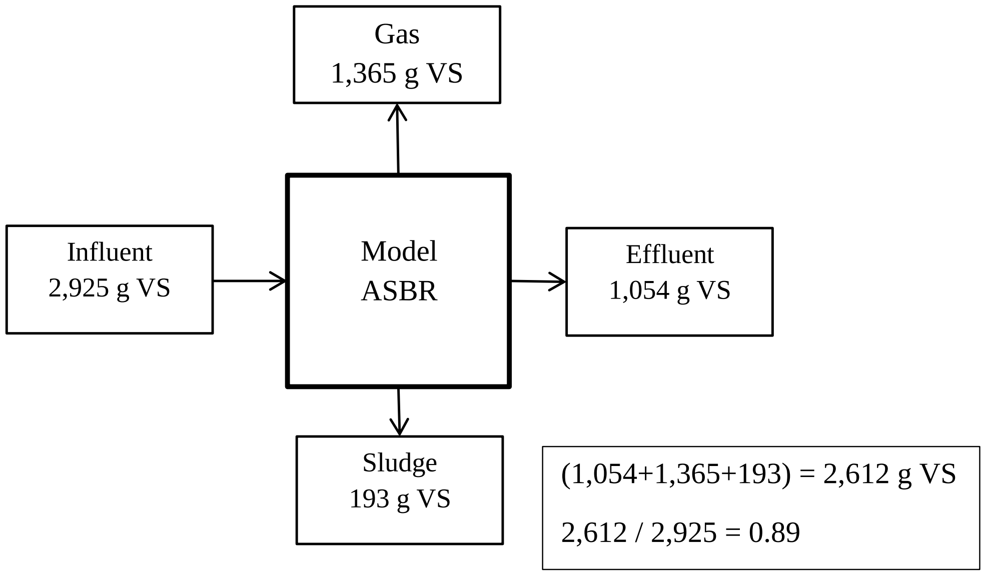

Figure 32. Volatile solids mass balance for glycerol treatment reactor

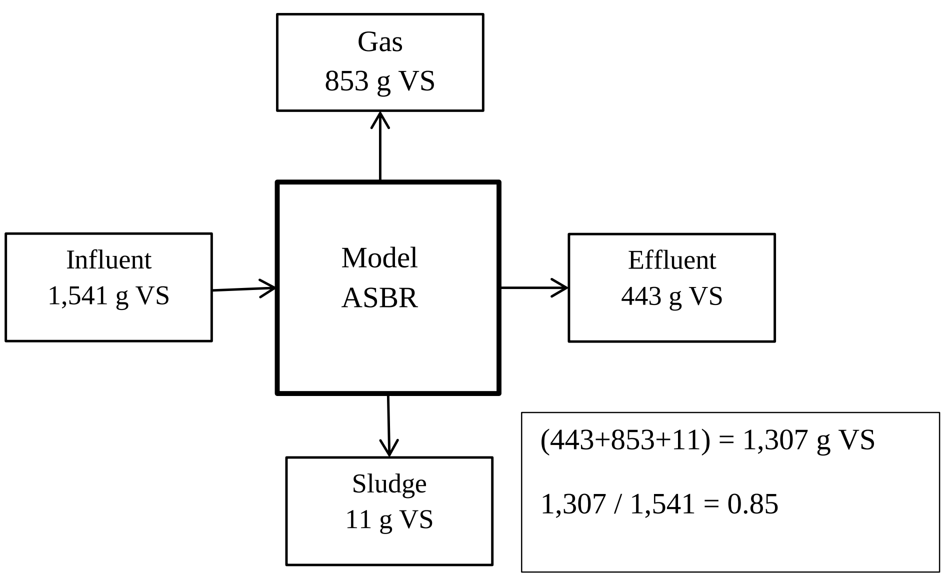

Figure 33. Volatile solids mass balance for control reactor

For the glycerol treatment reactor there are 313 g VS of influent not accounted for in the effluent, sludge, gas production. With a reactor operating volume of 18 l, this results in a mixed liquor volatile solids concentration of 17.4 g VS/ l. With an effluent VS/TS ratio of 0.71, the mixed liquor total solids concentration would be estimated at 24.5 g TS /l. For the control reactor, the 234 g of the influent volatile solids are not accounted for in the effluent, sludge and gas production. At a VS/TS ratio of 0.6 this yields mixed liquor solids concentrations of 13 g VS /l and 21.7 g TS / l.

During the steady state operation utilized for construction of the mass balance for the glycerol treatment and control reactor shows a higher effluent concentration for the treatment reactor. Operation at the 1% glycerol loading rate had previously shown that the treatment ASBR was capable of maintaining relatively equal effluent concentrations as the control reactor. Review of the data from the mass balance indicates that change in the timing of sludge wasting is required to maintain effluent concentrations due to increased biomass production from increased loading rates. The difference in total and volatile solid effluent mass between the treatment and control reactor during the 57 day period for that included two sludge wasting events are 862 g TS and 715 g VS. This results in a solids production increase of 15 g TS / day and 12.5 g VS / day for the glycerol treatment reactor.

(Glycerol Reactor 1,415 g TS) – (Control Reactor 553 g TS) = 862 g TS

(Glycerol Reactor 1,002 g VS) – (Control Reactor 287 g VS) = 715 g VS

It is assumed that the maximum mixed liquor solids concentration in the reactors is 20 g TS / l based upon the unaccounted solids (Influent – Effluent – Gas – Sludge = Unaccounted solids). Once the maximum mixed liquor solids concentration is reached in the reactor solids are no longer settled and retained within the reactor and leave via the effluent. Sludge wasting provides the ability to maintain the reactor solids concentration at or below the maximum mixed liquor solids concentration and provide both reactors with similar solids effluent concentrations given adequate digestion time. From the data collected during the 57 day period, it appears that sludge wasting was not performed adequately to manage the mixed liquor solids concentration of the glycerol treatment reactor such to provide effluent solids concentration similar to that of the control reactor.

If the reactor mixed liquor solids concentration is 20 g TS l^{-1} at the time of each sludge wasting event and the solids concentration is reduced to 15 g TS l^{-1}, 90 g TS are removed. A sludge wasting event of 90 g TS is within range of those observed during this period. With a TS production of 15 g / day this would require a sludge wasting event every 6 days. Assuming a 0.7 ratio of VS/TS, the period between sludge wasting for volatile solids would be 5 days based upon VS concentration. Thus a sludge wasting event for the glycerol treatment reactor operating at a 1% glycerol loading rate should occur every 5 to 6 days to maintain reactor effluent concentrations.

862 g TS / 57 days = 15 g day^{-1} TS production

715 g VS / 57 days = 12.5 g VS day^{-1}

(90 g TS [sludge]) / (15 g TS day^{-1} [production]) = 6 days

((90 g TS [sludge]) x 0.7 VS/TS) / (12.5 g TS day^{-1} [production]) = 5 days

The implementation of a sludge wasting every 6 days would have yielded 9 wasting events during the 57 day period. This would have yielded a total of 810 g TS and 567 g VS as sludge and an estimated effluent VS mass of 487 g. The changes to the sludge VS mass would decrease the reactor effluent VS mass within 44 g of the measured control reactor effluent; approximately a 10% difference between the control and treatment VS effluent mass (figure 34).

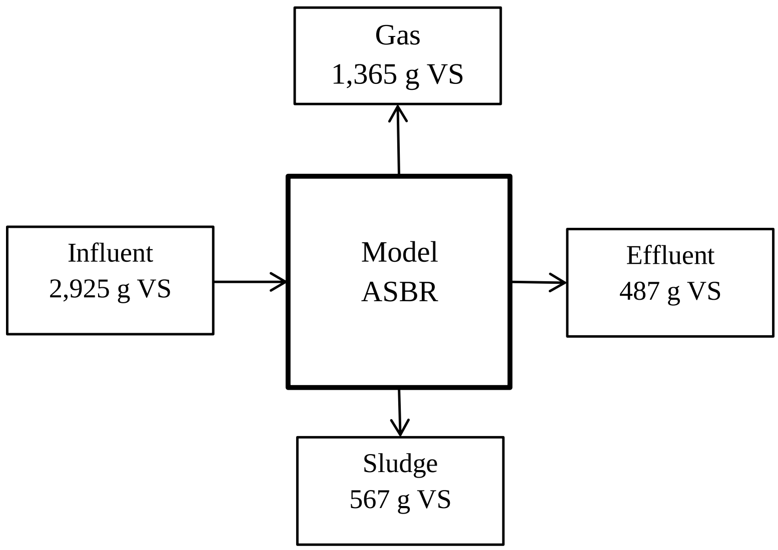

Figure 34. Volatile solids mass balance based upon theoretical 6 day sludge wasting frequency for glycerol treatment reactor for thesteady state period.

A nutrient mass balance for both reactors was completed for a period from January 21, 2012 to April 23, 2012, 84 days. From influent, effluent and sludge samples collected during this period the total nitrogen and phosphorus mass balance was constructed (figure 35 and 36). Sludge nutrient content was utilized for estimation of ASBR mixed liquor nutrient content as sludge samples were taken after the react phase while mixing was occurring, providing an approximation of the mixed liquor contents.

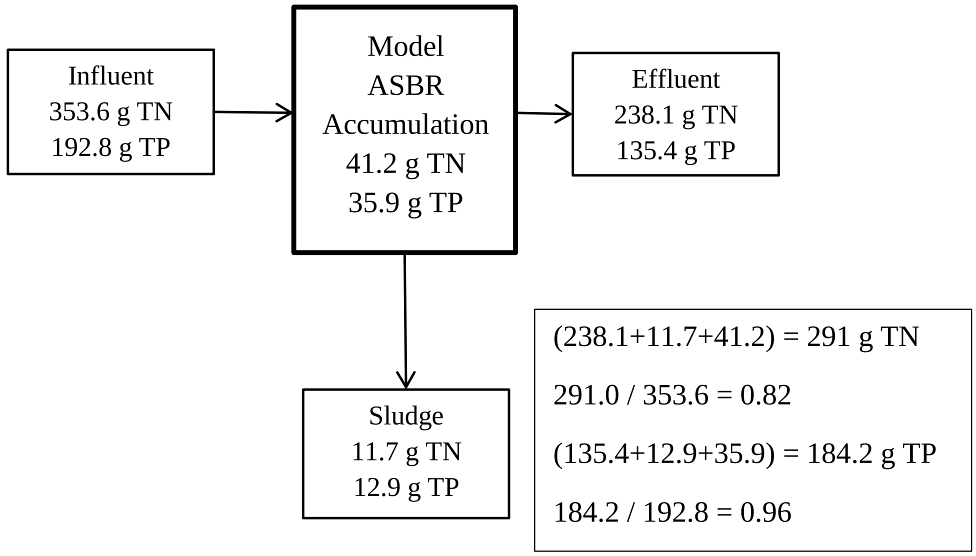

Figure 35. Total Nitrogen and Phosphorus mass balance for glycerol treatment reactor

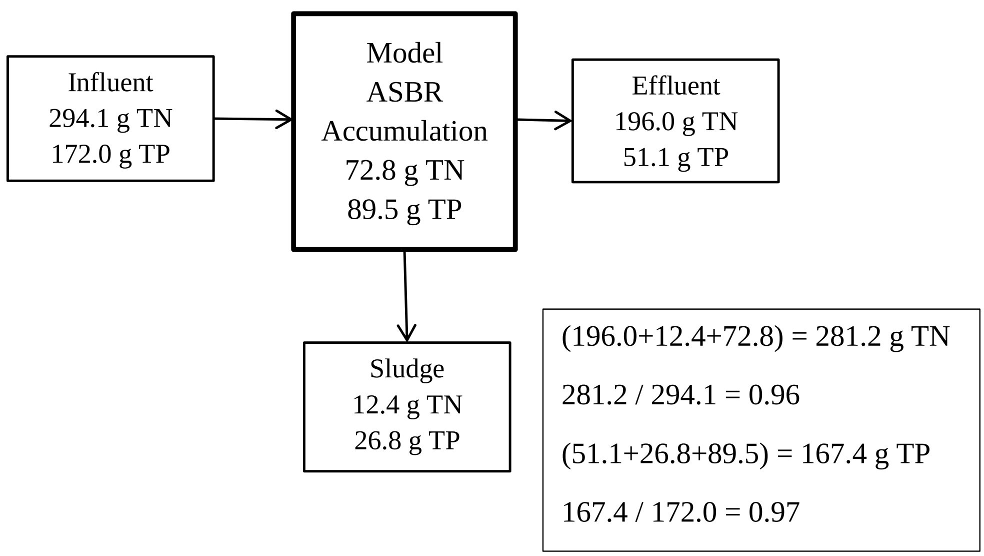

Figure 36. Total Nitrogen and Phosphorus mass balance for control reactor

Nitrogen losses in the biogas were estimated based upon biogas quality and production. The estimated nitrogen mass emitted from biogas is 51.74 g and 5.40 g N for the glycerol and control reactor respectively. The inclusion of the biogas nitrogen mass increased the mass of influent nitrogen accounted for from 82% to 97 % for the glycerol treatment reactor. For the control reactor the mass of influent nitrogen accounted for including the biogas nitrogen increased from 96 to 97%. The significant increase in gaseous nitrogen losses from the co-digestion of the crude glycerol is a direct result of the increased biogas production.

Conclusions

Crude glycerol is highly digestible when added as a co-digestion feedstock to ASBR digesters. The combination of high digestibility and energy density makes crude glycerol an ideal candidate for co-digestion in an ASBR. The high energy density allows low volumetric inclusion rates to be utilized without significant operational changes to new and existing ASBR reactors. The maximum inclusion rate of glycerol to an ASBR treating low strength swine manure (<1% TS) was 1.1% of influent volume. Stable operation is maintained at an inclusion rate of 1.0%. At 1.0% inclusion rate of crude glycerol, methane production increases of 600 to 800% are expected for an ASBR treating low strength swine manure. COD removals of 80% at 1.0% inclusion rate are also achieved. Even at the higher removal rate, effluent organic matter concentration should increase slightly as glycerol is added. It may be possible to maintain effluent quality of a glycerol treated digester equal to a manure only digester by altering sludge wasting rates. Glycerol co-digestion in an ASBR treating low strength swine manure will not have a significant impact of effluent nutrient quality. The one expected change in the nutrient mas balance is increase loss of volatile nitrogen in the biogas due to the significant increase in total biogas production due to glycerol inclusion.

CHAPTER V

CONSIDERATIONS FOR DESIGN OF ANAEROBIC SEQUENCIGN BATCH REACTORS TREATING SWINE EFFLUENT

Introduction

The biological treatment of aqueous organic waste streams consists of two pathways; aerobic and anaerobic. The biological conversion of energy contained within the waste stream distinguishes the aerobic pathway as a waste treatment process and anaerobic as waste treatment and energy production process. During the aerobic removal of waste stream Chemical Oxygen Demand (COD), energy, is converted to biomass, CO_2, and H_2O. While the anaerobic conversion of waste stream COD produces recoverable energy in the form of CH_4 in addition to biomass and CO_2. The biomass production during aerobic treatment is 6 to 8 times that of anaerobic digestion and energy intensive due to the oxygenation of the treatment system (Metcalf and Eddy, 2003). The production of CH_4 and lack of input energy for anaerobic treatment results in a net energy production of nearly 7 times that of the input energy required for aerobic treatment as shown below:

Assumptions (Metcalf and Eddy, 2003):

- 0.8 kg oxygen required for removal of 1 kg COD

- 1.52 kg oxygen per kW-hr aeration efficiency
- 0.35 m^3 CH_4 kg COD^{-1} removed methane production rate
- 9.96 kW-hr m^{-3} methane energy content (0° C and 1 atm)

Aerobic removal of 1 kg COD

0.8 kg O_2 kg COD^{-1} removed x kW-hr 1.52 kg O_2^{-1} = **-0.52 kW-hr kg COD removed^{-1}**

Anaerobic removal of 1 kg COD

0.35 m^3 CH_4 kg COD removed^{-1} x 9.96 kW-hr / m^{-3} CH_4 = **3.49 kW-hr kg COD removed^{-1}**

For an anaerobic digestion reactor to provide net positive energy rates the operational energy input must total less than 3.49 kW-hr for each kg of COD removed. This, however, is for theoretical operation. Using specific biomass production rates for known reactor types and waste streams, the maximum inputs can be determined for positive net energy production. For laboratory scale ASBR models at varying HRT's, operating temperatures, and substrates the average specific methane yield is 0.2 l CH_4 g COD^{-1} with a range of 0.1 to 0.32 l CH_4 g COD loaded^{-1} (Table 17). In terms of energy production this equates to 1.0 to 3.2 kW-hr as methane per kg COD loaded. The shaded portion of figure 37 illustrates the potential methane energy production rates based upon ASBR methane yields for OLR's up to 12 kg COD m^{-3} day^{-1}.

Table 17. Laboratory scale ASBR operational parameters, influent substrate, and methane production

HRT	Temp	OLR		Average Specific Methane Yield	
Days	°C	g COD / l-day	Waste Stream	l CH4 / g COD loaded	Source
1.5-10	35	0.4-9.4	Landfill leachate	0.21	Timur and Ozturk, 1999
0.5-2	35	2.0-12.0	Non fat dry milk	0.31	Sung and Dague, 1995
2-6	35	0.8-5.5	Swine manure	0.31	Zhang et al., 1997*
1.25-5	35	1.5-12	Synethitic	0.17	Cheong and Hansen, 2008
10-20	35	2.71-5.42	Thermally hydrolyzed sewage sludge	0.24	Wang et al., 2009
1	33	1.5-5.0	Brewery Wastewater	0.32	Shao et al., 2008
0.83	30	1.15-4.79	Whey	0.1	Mockaitis et al., 2006
5.25	20	0.8	Swine Manure	0.14	Ndegwa et al., 2005
6	35	0.8	Swine Manure	0.16	Ndegwa et al., 2005

*– OLR and Average Specific Methane Yield in term of Volatile solids; g VS / l-day and l CH4 / g VS $_{loaded}$

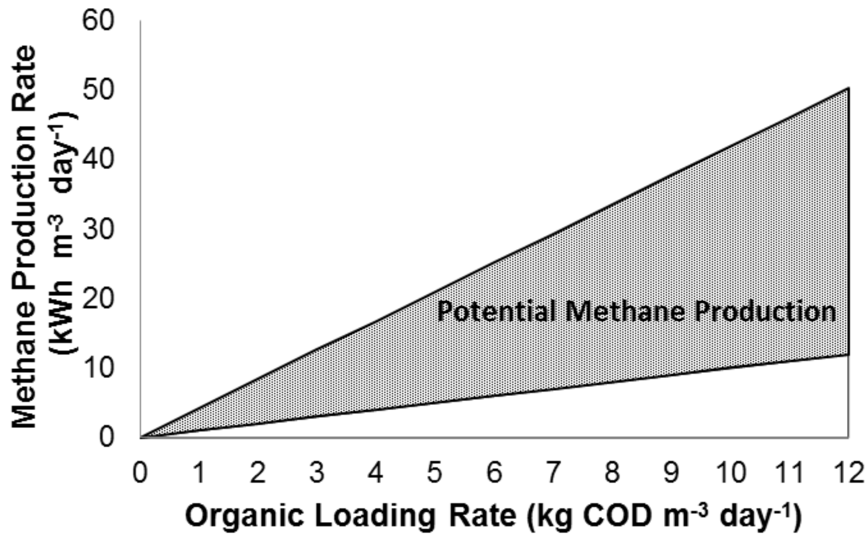

Figure 37. Potential methane energy for laboratory scale ASBR reactors.

The biological methane production potential, loading rates, and mechanical energy inputs must all be considered in the design of ASBR for bioenergy production. In some locations and industries, the need for excess electrical energy may not be required due to local net metering regulations. Net metering provides for the buyback or crediting to the facility of excess electrical generation by the utility provider. However, during the design process it is impossible to forecast changes to facility infrastructure, operational energy requirements, or utility net metering regulations, thus the most efficient design within the limits of practicality and feasibility should be utilized.

Objectives

- Development of design steps for consideration during the design of a full scale ASBR for treatment of low strength swine manure

Design Steps

The four primary operational energy inputs for an ASBR are influent transfer, mixing, heating, and effluent transfer. The energy demands of influent and effluent transfer and heating are a function of the reactor's site location. Mixing is a function of mixing type, mixed liquor solids concentration, reactor size, and geometry. The first four steps in the design process will provide the information necessary to determine the potential for positive energy production from influent waste stream. This process outlined in figure 38, is based upon the waste stream's potential to meet the energy requirements of influent heating.

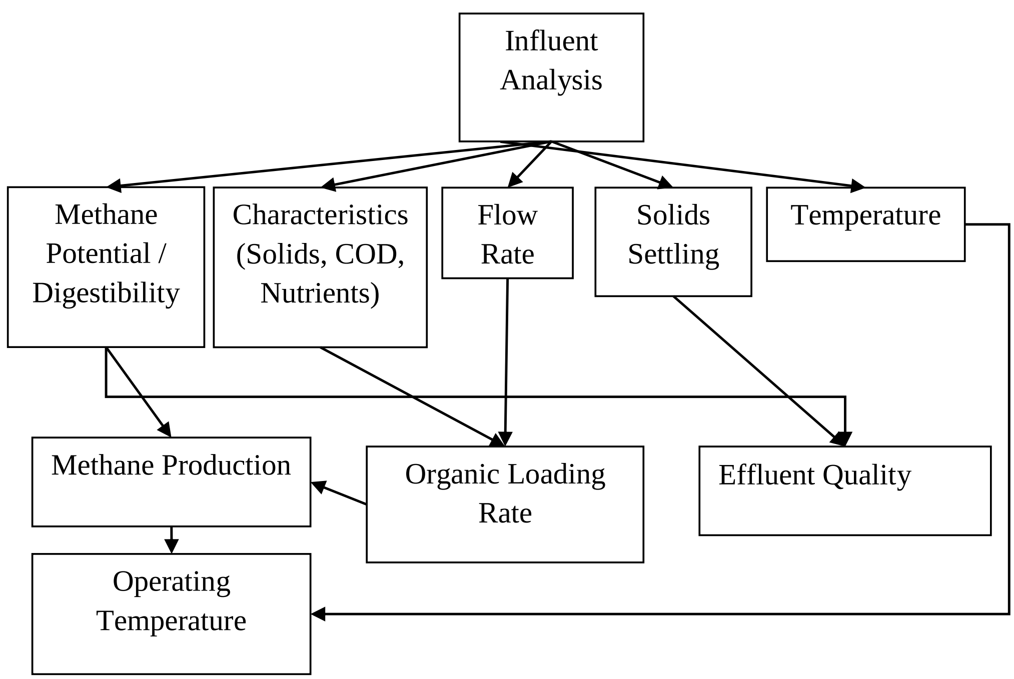

Figure 38 – ASBR design process flow chart

Influent Analysis

Measurements of five influent parameters are required. The first two parameters are the measurement of the digestibility or methane potential and the characteristics of the influent. The methane potential and digestibility of the waste stream can be accomplished utilizing several methods including, laboratory scale models, biological methane potential test (BMP), and toxicity tests. Completion of BMP's and toxicity tests at the expected operating temperature provides the potential methane yield and loading rate.

The solids profile, COD, nutrient analysis, and flow rate will provide the necessary data for determination of the potential organic loading rates and loading rates of necessary nutrients and potential toxicants. The influent pH and alkalinity give an initial estimation of the need for supplemental alkalinity. Data regarding the influent temperature is required for the design of the reactor operational temperature and heating system components. The influent temperature along with methane production potential of the waste stream will be used to determine the required organic loading rate required to meet the energy demand for influent heating. Solids settling analysis of the influent provides an initial estimation of the mixed liquor solids settling velocities. The analysis of the influent solids settling should provide a relationship between settling velocity and solids concentration. This relationship will be utilized to determine the design mixed liquor solids concentration which will be used for determination of the settling phase length, mixing intensity, and sludge wasting period.

Organic Loading Rate

The volumetric organic loading rate, OLR, is the ratio of the organic mass entering the reactor per day to total reactor volume (eq. 14). The OLR is expressed in terms of either Volatile Solids (VS) or COD with units of kg VS m^{-3} day^{-1} or kg COD / m^{-3} day^{-1} and is a function of the hydraulic retention time (HRT) and influent concentration. As shown in table 16 the OLR loading rates of laboratory scale ASBR's ranges from 0.4 to 12 kg COD m^{-3} day^{-1}.

For waste streams concentrations resulting in high or low OLR's, two methods can be employed to modify the OLR. The first option is lengthening or shortening the HRT, the second is dilution or concentration of the influent waste stream. There are operational and design considerations that must be considered prior to implementation. Dilution of the waste stream and increasing the HRT results in increased reactor volume, increasing construction costs and potential heating and mixing inputs. Reduction of the HRT and concentration of the influent waste stream can result in incomplete biological utilization from overloading and increases the potential for solids washout.

Methane Production

There are two expressions of methane production; methane yield and volumetric methane production rate. The methane yield is the ratio of methane produced per mass of organic material loaded in the reactor, expressed as either m^3 CH_4 kg VS^{-1} or m^3 CH_4 kg COD^{-1}. The volumetric production rate is the volumetric ratio of methane produced per reactor volume per day, expressed as m^3 CH_4 m^{-3} day^{-1}. The volumetric production rate is function of the methane yield at a given OLR and temperature. The volumetric production rate can be estimated for the ASBR using a modified form of Chen and

Hashimoto's Contois equation (eq. 11, 12 and 13) (Chen and Hashimoto, 1980 and Hashimoto, 1983).

Operating Temperature

A benefit of the ASBR's ability to retain and accumulate active biomass within the reactor is the option for operational temperatures less than 35°C (Dague et al., 1998). Laboratory ASBRs operated in the mesophilic and psychrophilic ranges have achieved COD removals of 75% or higher (Dague et al., 1998; Ndegwa et al., 2005; and Ndegwa et al., 2008). The ability to utilize reduced operating temperature provides an input energy reduction of 1.16 kW-hr per m^3 of influent for each 1˚C drop in temperature given no heat loss from the system and an influent temperature is less than the operating temperature. The reduction of the operating temperature from 35˚C to 20°C as examined by Ndegwa reduces the daily influent heating requirement by 17.4 kW-hr m^{-3}.

When considering the design operational temperature, the average influent temperature and OLR are required to estimate the influent heating energy balance. At a 5 day HRT, the influent heating requirement, given a thermal conversion efficiency of 41.5% and 1°C temperature difference the required OLR for methane yields of 0.1 and 0.32 m^3 CH_4 kg COD^{-1} are 0.56 and 0.17 kg COD m^{-3} day^{-1} (Dresser-Rand, 2013). The required OLR at both methane yields for influent-effluent temperature differences of 1, 5, 10, and 15 °C are given in table 18. The incorporation of a Combine Heat Power unit for the conversion of biogas to electrical power and recovery of waste heat from internal combustion engine allows for the electrical and thermal energy inputs to be recovered from a single source.

The Theoretical Influent Heating Requirement (kW-hr m^{-3} reactor volume-day) is first calculated:

$$TIHR = (T_R - T_I) * (1.16 \text{ kW-hr} / m^{-3}\ °C^{-1}) / (\Theta_H) \qquad (21)$$

where:

T_I= influent temperature (°C)

T_R= Reactor temperature (°C)

TIHR= Theoretical Influent Heating Requirement (kWh m^{-3} reactor volume-day)

Given the THIR the Design Influent Heating Requirement (kW-hr m^{-3} reactor volume-day) is determined:

$$DIHR = (THIR) / (CHP_H) / 100 = \qquad (22)$$

where:

CHP_H – Combined Heat Power Heat Recovery Efficiency (%)

DIHR -Design Influent Heating Requirement (kW-hr m^{-3} reactor volume-day)

The required methane production volume for influent heating is then calculated:

$$HMPV = (DIHR) / (9.96 \text{ kw-hr } m^{-3}\ CH_4) = \qquad (23)$$

where:

HMPV= Influent Heating Required Methane Production Volume (m^3 CH_4 m^{-3} day^{-1})

Given the HMPV the required OLR for influent heating is calculated:

$$OLR_H = (HMPV) / (SMY) \qquad (24)$$

where:

OLR_H= OLR for Influent Heating (kg COD m^{-3} day^{-1})

SMY= Specific Methane Yield (m^3 CH_4 kg COD_{loaded}^{-1})

Table 18. Organic loading rate requirements for influent heating for design heat transfers, effect of temperature and SMY

Influent-Effluent Temperature Difference	Heating Input	Required OLR at 0.1 m^3 CH_4 kg COD $_{loaded}^{-1}$	Required OLR at 0.32 m^3 CH_4 kg COD $_{loaded}^{-1}$
°C	kWh m^{-3} day^{-1}*	kg COD l^{-1} day^{-1}	kg COD l^{-1} day^{-1}
1	0.23	0.56	0.17
5	1.16	2.81	0.88
10	2.34	5.64	1.76
15	3.49	8.44	2.64

*Total reactor volume

Influent and Effluent Transfer

The ideal site location of an ASBR system provides adequate slope between the influent source, reactor and effluent storage to allow for gravity transfer. As the ASBR is a batch process, batch influent flows are required at the same time as the feed phase in addition to adequate flow rate. For influent sources and locations where batch influent flow and/or gravity flow are not available, additional reactor system components are required. The design of components for transfer of influent and effluent to and from the reactor is straight forward requiring only that the batch volume can be transferred to and from the reactor during the feed and decant phases. The feed phase, although a separate phase, is potentially part of the react phase.

Two laboratory scale experiments reviewed the impact of the ratio of feed to react phase length with regard to reactor performance. Although both studies utilized a synthetic influent, both indicated similar results for reactor total volatile fatty acid (TVA) concentrations during the react phase. The first study utilized a low strength influent, 0.5 g COD / l with an OLR of 0.8 g COD / l-day and feed to react ratios of 0.2 to 0.97 (Rodrigues et al., 2003). At varying feed to react ratios, no significant difference was found for effluent COD and suspended solids. Although no difference was observed for effluent TVA concentrations, the peak TVA concentrations during the react phase were reduced by 25% when the feed to react ratio was increased above 0.73. The second study reviewed had higher OLR's, 1.5 to12 g COD / l-day, and influent concentrations, 7.5 to 30 g COD / l at HRT's of 1.25, 2.5 and 5 days (Cheong and Hansen, 2008). The feed to react ratio ranged from 0.01 to 0.83. It was observed that by increasing the feed to react phase length, there was an increase in the specific methane yield which is attributed to reduced peak TVA concentrations during the react phase. Although these studies do not provide a design recommendation as to the optimum ratio of feed to react phase length, consideration should be given during the design process. Considerations should include the ability to modify the influent flow rates as an option for physical control of the TVA concentrations during the react phase, aiding in reactor operational stability and performance.

The decant phase length is related to the settling phase length and settling velocity of the mixed liquor solids. The length of the settling phase is the time required for the mixed liquor solids to settle under zone settling conditions past the effluent withdrawal point in the reactor. During the decant phase, no mixing occurs and the solids continue to settle. The decant phase length must not exceed the time required for the compression settling to begin as this increases the mixing intensity required to resuspended the settled solids. As the zone settling velocity is a function of the influent substrate and mixed liquor solids concentration, the maximum decant phase length is unique to each reactor. As little to no biological treatment occurs during the settling and decant phase, all effort should be made to minimize the length of these phases.

The steps for determination of OLR required to meet the influent and/or effluent transfer pumping energy requirements as follows:

Calculation of the transfer pump power:

$$W_P = H \times Q \quad (25)$$

Where:

W_P – Pump Power, kW

Q – Transfer flow rate, $m^3\,s^{-1}$

H – Required pump pressure head, kPa

The transfer flow rate is calculated as follows:

$$Q = V_c / t_F / 3600 \quad (26)$$

V_c – cycle volume, m^3

t_F – transfer phase length, hr

Calculation of the transfer pump brake horse power:

$$W_B = W_P / (P_{eff}) / (D_{eff}) \quad (27)$$

W_B – Brake Pump Power, kW

P_{eff} – Pump Efficiency, frac.

D_{eff} – Motor Drive efficiency, frac.

Calculation of the transfer pump daily energy requirement:

$$E_{pump} = W_B \text{ x } t_F \text{ x } R \tag{28}$$

Where:

E_{pump} = Pump energy requirement, kw-hr day^{-1}

Calculation of the reactor transfer pumping requirement (RPR) kWh m^{-3} day^{-1}:

$$RPR = \Sigma\, E_{pump} / V_c / HRT / R \tag{29}$$

Calculation of the design reactor transfer pumping requirement (DRPR), kWh / m^{-1} day^{-1}:

$$DRPR = (RPR) / (CHP_E) / 100 \tag{30}$$

Where:

CHP_E= Combine heat and power system electrical energy recovery, %

Calculation of the daily methane production volume required for electrical energy requirement of the transfer pumps:

$$PMPV = (DRPR) / 9.96 \text{ (kWh } m^{-3} \text{ } CH_4) \tag{31}$$

where:

PMPV – Transfer pumping methane production volume, m^3 CH_4 m^{-3} day^{-1}

Calculation of the OLR required to meet the electrical energy requirements for transfer pumping:

$$OLR_P = (PMPV) / (SMY) \tag{32}$$

Where:

OLR_P – OLR for reactor transfer pumping, (kg COD m^{-3} day^{-1})

Given a 1,000 m^3 reactor (12 m diameter and 9 m height) operating at 2 cycles per day and a 5 day HRT, the OLR required for meeting the influent and effluent energy requirements is outlined in table 19. The assumed discharge head for the transfer pumps is 138 kPa (20 psi) with an influent elevation head of 89.7 kPa (30 ft) and effluent elevation head of 44.8 kPa (15 ft). The elevation heads assume that influent must be pumped from a below grade lift station into the reactor and reactor effluent must be pumped over the waste storage structure embankment. For this reactor the OLR and influent concentration required to meet the influent and effluent transfer pumping requirements would be 0.051 kg COD m^{-3} day^{-1} 0.25 kg COD m^{-3}, respectively.

Table 19. Example calculation for required OLR for meeting influent and effluent transfer pumping requirements

Parameter	Influent	Effluent
Cycle Volume, V_c (m^3)	100	100
Feed Phase Length, t_F (hr)	0.25	-
Decant Phase Length, t_D (hr)	-	0.25
Flow Rate, Q, ($m^3 s^{-1}$)	0.11	0.11
Head, H (kPa)	228	183
Pump Power, W_P (kW)	25.1	20.1
Pump Efficiency, P_{eff} (frac.)	0.75	0.75
Motor Drive Efficiency, D_{eff} (frac.)	1.0	1.0
Brake Pump Power, WB (kW)	33.5	26.8
Pump Energy Requirement (kWh day^{-1})	16.8	13.4
Reactor Transfer Pumping Requirement, RPR (kWh m^{-3} day^{-1})	0.0168	0.0134
Combine Heat and Power Electrical Efficiency, CHP_{eff} (%)	30	30
Design Reactor Transfer Pumping Requirement, DRPR (kWh m^{-3} day^{-1})	0.056	0.045
Transfer Pumping Methane Production Volume, PMPV (m^3 CH_4 m^{-3} day^{-1})	0.0056	0.0045
Specific Methane Yield, SMY (m^3 CH_4 kg COD_{loaded}^{-1})	0.2	0.2
OLR for Reactor Transfer Pumping, OLRP (kg COD m^{-3} day^{-1})	0.028	0.023
Total OLR required for Influent and Effluent Transfer, (kg COD m^{-3} day^{-1})	0.051	

Influent Flow Control

The batch operation of the ASBR will typically require the inclusion of an influent storage vessel to buffer influent flow and provide cycle treatment volumes at the require times. In addition to converting continuous or semi continuous influent flows to batch volumes, the vessel provides the opportunity for managing changes to influent characteristics during daily operations. The size of the buffering vessel should be at least 1 HRT in order to provide for operation of the ASBR at 1 cycle per day. Increased vessel capacity should be considered based upon influent flow variations. For example if the facility is only in operation during the weekdays, a buffering vessel volume of 3 HRT's would be recommended to maintain ASBR operation during weekends and holidays. Design of the buffering vessel should include a mixing component; however continuous mixing of the vessel is not required. Mixing time and intensity should be such that the stored influent solids can be rapidly suspended and any scum layer can be reincorporated. Controls for the mixing component should be provided to initiate mixing prior to and during the feed phase and terminate after the feed phase has been completed.

For influents such as manure and unscreened municipal or domestic waste waters a screening component or grinder pump should be installed as part of the influent transfer system. A course screening of influents such as manure and domestic waste waters will provides the opportunity to remove trash and other non-digestible as well as removing large solids that may cause damage to pumps or clogging of the pipes. A grinder pump for transfer of the influent waste stream can reducing large particles prior to digestion increasing the net surface area. During the design of the screening or particle size reduction components, the accessibility for cleaning and maintenance of these components must be considered. For example the installation within a lift station can create a confined space entry if the screen or pump cannot be readily removed for maintenance and cleaning.

Mixing System

Mixing of an anaerobic digester, regardless of mode, has five desired design results; temperature maintenance, substrate distribution, sedimentation prevention, scum and crust prevention and release of entrapped gases (Mills, 1979; Ward et al., 2008). For the ASBR, sedimentation prevention is not a priority in the design of the mixing system as settling of the mixed liquor solids is a key parameter to successful ASBR operation. Thus, off the bottom rather than complete mixing can be utilized, reducing the overall energy requirements for operation. However, adequate mixing for the release of entrapped gases is necessary to prevent the flotation of granules and solids that may hinder the settling ability of the mixed liquor during the settling phase.

A wide range of applied and recommend mixing intensities are presented in the literature (figure 11). The range of applied mixing intensities shown in figure 11 range from the USEPA's recommend 5 – 8 W m^{-3} to 3,500 W m^{-3} utilized by Bhutada and Pangarkar (USEPA, 1979 and Bhutada and Pangarkar, 1989). Of the literature reviewed, the experiments conducted by Karim et al. (2005) from a design standpoint provide the most useful insight as to the needs of reactor mixing. Karim et al. (2005) operate four reactors, each with separate mixing regimes; unmixed, biogas recirculation, mixed liquor recirculation and impeller mixed. The three mixed reactors were each mixed with an applied mixing intensity of 8 W m^{-3}. The effect of mixing or lack of mixing, mixing regime and influent concentration were examined. Karim et al. (2005) found that at influent solids concentrations of 5% there was no difference in reactor performance and methane production. At the 10% influent solids concentration, the methane production for the unmixed reactor was significantly reduced compared to he mixed reactors. For the mixed reactors at both concentrations, there was no significant difference between the three mixing regimes.

In laboratory scale ASBR models, all three of the typical mixing regimes have been utilized; impeller, biogas recirculation, and mixed liquor recirculation (jet mixing) (table 2). In full scale application, only impeller and jet mixing have been utilized, although only two full scale reactors have been constructed and operated. The Iowa State University utilized two 3 kW Flygt 4600 series submersible impeller mixers (Angenent et al., 2002). The Oklahoma State University ASBR utilized a three nozzle US Filter model 80 connected to Fairbanks Morse centrifugal turbine pump with an 11kW motor and 69 l /s flow rate.

The design mixing intensity and power consumption of the OSU ASBR mixing system was 14.6 W m^{-3} and 253 kWh day^{-1}. During the continuous operation of the OSU ASBR, it was found that reducing the flow rate to the jet mixing pod to 9 l s^{-1} provided adequate mixing without reducing the performance of the ASBR (Steele and Hamilton, 2009; Steele and Hamilton, 2010). The reduction of the jet mixer flow rate reduced the mixing system power consumption to 129 kWh day^{-1}, a 124 kWh day^{-1} reduction.

Utilizing the ratio of design mixing intensity of the OSU ASBR of 14.6 W m^{-3} and daily power requirement of 253 kWh day^{-1} daily power requirement and volumetric input power requirement are estimated. This yields a daily input power requirement of 139 kWh day^{-1} and volumetric power requirement of 0.34 kWh m^{-3} day^{-1} at a mixing intensity of 8 W m^{-3} similar to power input requirements of the reducing mixing intensity of the OSU SREC ASBR (Steele and Hamilton, 2009; Steele and Hamilton, 2010).

Based upon the recommendation of the USEPA and the results of Karim et al., (2005a and 2005b) and Steele and Hamilton, (2009 and 2010), an initial design applied mixing intensity of approximately 8 W m^{-3} should be utilized. The daily volumetric mixing power requirement can then be converted the required OLR for meeting the electrical power requirements of the mixing system. Given a CHP electrical conversion efficiency of 30%, a SMY of 0.2 m^3 CH4 kg COD_{loaded}^{-1} and methane energy value of 9.96 kWh m^{-3}, the required OLR is 0.57 kg COD m^{-3} day^{-1}.

Automated Control System

An automated control system for an ASBR at a minimum requires four input variables, react phase length, settling phase length, react phase mixed liquor depth/height and decant phase mixed liquor height.

Reactor Volume Control

Real time measurement of the influent, effluent, and reactor volume are required for both manual and automated operation of the ASBR. For a continuous flow reactor the reactor outfall maintains a constant reactor volume and allows effluent flow to equalize to the influent flow rate. The ASBR batch process requires measurement of both the influent and effluent volumes during each batch to maintain a constant volume and HRT. There are two sensory methods for measurement of reactor flows and volumes, volumetric measurement of influent and effluent flows and depth measurement of the reactor vessel. Reactor depth measurement of the ASBR is the recommended choice for reactor flow and volume control, as measurements can be readily verified and mixed liquor height to effluent withdraw outlet is key to solids separation and reactor performance.

For a fixed flexible membrane-covered ASBR like that utilized by Iowa State University and Oklahoma State University, an exterior standpipe equipped with a narrow beam ultrasonic level senor should be utilized. The OSU SREC ASBR, as shown in figure 19, is equipped with a narrow beam ultrasonic level senor for control of the mixed liquor height within the reactor. Using the changes in mixed liquor height, the influent and effluent volumes are controlled during fill and decant phases. The low pressure biogas pressure of the flexible membrane cover, approximately 5 to 7.6 cm H_2O, eliminates the need to calibrate the level sensor to the internal reactor pressure. For other cover types calibration of the level sensor to the internal biogas pressure and changes to installation location are required.

Solid fixed reactor covers can allow for biogas pressures of 24” W.C., requiring standpipe ultrasonic level senor measurements to be compensated to accurately reflect the mixed liquor height. To reduce the need for measurement of the biogas pressure for level sensor calibration, when installed in an external standpipe the installation location can be changed. Removing the level sensor from the external standpipe and it mounting through the fixed solid cover, no pressure compensation is required for mixed liquor height measurement. For sensor height verification, internal biogas pressure would be required for comparison of senor reading to external standpipe heights.

During the operation of the OSU SREC ASBR, several failures of ultrasonic level sensor were observed due to condensation on the sensor due to its external location. The water vapor from the warm mixed liquor in the exterior standpipe would condense on the sensor during cold weather, producing false high liquid level readings. The result of these high mixed liquor height readings was the unnecessary release of mixed liquor from the reactor. The PLC programming's interpretation of the false high readings is an overfilled reactor and opens the effluent control valve, releasing mixed liquor until the sensor readings are within the set height parameters. During the programming of the PLC, programming and installation of the ultrasonic level senor for use with an external standpipe steps for prevention of this scenario should be considered.

Basic Control Steps

The programming controls for automated control of the ASBR are based upon the four phases of operation and a standby mode. The standby mode provides a holding phase in the event operational alarms are triggered or for maintenance of reactor components. A basic description of the reactor phases and operations during each phase are as follows:

1. Fill – Prescribed influent volume or level is transferred into reactor or until operating mixed liquor volume / level reached.
2. React - Mixing system turned on and operated for the prescribed react phase length
3. Settle – Mixing system turned off and settling allowed for prescribed phase length
4. Decant - Prescribed effluent volume/level is transferred from the reactor

5. Standby – Additional phase to be included to act as a holding phase as a result of a feedback alarm or for temporarily placing the reactor in a static state. Mixing system remains operational at reduced capacity for prevention of compression settling and temperature maintenance.

Automated Operational Feedbacks

Based upon the initial programming and operational observations of the OSU SREC ASBR four control feedbacks are recommended for inclusion in control programming. The operational feedback inputs are low influent volume, low reactor volume, high reactor volume and pump failure. The inclusion of these operation input variables in the automation programming will aid in reducing energy consumption and unnecessary pump wear.

Low Influent Volume

The system's ability to acknowledge the lack of influent will prevent the operation of influent feed pump without adequate liquid, preventing damage. Additionally, this will allow the reactor to go into standby until adequate influent volume is available for operation. Placement of the reactor into standby mode reduces the energy consumption of the reactor to only that required for temperature maintenance and prevention of compression settling. If the reactor is not placed into standby, the fill phase would continue until the operation volume/level was reached, allowing the mixing system to continue to operate at full capacity.

Low Reactor Volume

A low reactor volume/level feedback alarm aids in the prevention of the loss mixed liquor solids through the unintended released of mixed liquor due to electrical and/or mechanical failures. By insuring that a minimum volume of mixed liquor is retained within the reactor the time to restore reactor performance after any unintentional release of mixed liquor can be reduced. A separate mixed liquor level sensor should be utilized to verify the minimum level with the primary operational level sensor. This will help to provide a failsafe if the primary mixed liquor level sensor should fail. The control sequence in the event of a low level alarm should include the termination of all pumps and closing of reactor valves. Additionally, a manual valve on the sludge removal outlet line should be utilized to prevent the release due to failure of an automated sludge removal line valve.

High Reactor Volume/Level

From observations during the operation of the OSU SREC ASBR, the response to a high reactor volume reading should be considered. For the OSU ASBR, the response to the high reactor volume was to open the effluent valve and release the excess volume. However, for the each of the times that a high reactor volume alarm was triggered, the reactor volume was not in excess of the set high level, but rather a false reading of the ultrasonic level sensor occurred, thus resulting in the unintentional release and wasting of mixed liquor solids as the mixing pump was still in operation. The recommended response to the high reactor volume/level alarm should be as follows:

- Closing of influent piping valve(s)
- Turning off of reactor mixing and influent feed pumps

- 1 hour operational pause

If at the end of the 1 hour operational pause, the reactor volume is still above the high reactor volume/level set point, the reactor is placed into the standby phase. However, if the reactor volume/level is below the set point, reactor operation will resume at the point prior to the high reactor volume/level alarm.

Pump Failure

Although thermal switches may be installed on the pumps, this may not completely prevent damage to reactor pumps. Also, submersible pumps such as those used in a lift station may not be easily accessible for resetting of the thermal switch. Three options should be considered for pump failure protection; flow sensors, pressure sensors, and timers. A flow or pressure sensor may be placed after pumps to detect for high and low limits during pump operation. Additionally, a timer may be considered to allow for pumps to run only for predetermined maximum periods. However, a timer will not provide protection for the mixing system pump as this pump will run for extended periods during the react phase or its entirety.

Sampling Accessibility

A design consideration that was observed to be overlooked during operation of the OSU SREC ASBR was the accessibility for sampling of influent, effluent and mixed liquor. Readily accessible and safe points of sampling must be provided in the both influent and effluent transfer system. Considerations for the sampling locations include the ability for manual and automated sample collection. In addition to accessibility, location should provide for a well-mixed and homogenous sample that will be representative of the flow. For mixed liquor, two types of sampling locations are desired to aid in operational management. The first is a sampling of the mixed liquor for estimation of reactor mixed liquor solids mass. This is accomplished during the react phase after mixing has fully developed and fully suspended the mixed liquor solids. Sampling locations may be placed within the mixing system. The second sample type is for determination of solids stratification within the reactor. The ability to determine the stratification of solids in the mixed liquor during the react and settle phase provides operational feedback as to uniformity of mixing and effectiveness of settling. For solid reactor covers, both fixed and floating, this can be accomplished by the additional of a port with a conduit extending below the mixed liquor surface. The extension of the observation port by a conduit to below the mixed liquor surface provides a liquid-gas seal, reducing the introduction of atmospheric oxygen into the reactor during sampling. For flexible membrane covers this set up can also be installed along the perimeter of the cover which is accessible from exterior platform mounted on the reactor vessel.

Gas Control and Handling System

Two gas handling system problems that were encountered during the operation of the OSU SREC ASBR were condensation and gas meter failures. During cooler weather, water vapor in the biogas condensed in the flame/spark arrestors located upstream of the condensation collector. The large surface area of metallic fins contained within the arrestors provides an ideal location for vapor condensation. This required nearly daily drainage of the arrestors and insulation to prevent freezing. Freezing of the condensate restricts gas flow, causing biogas pressure buildup, triggering the biogas pressure relief valve and subsequent release of biogas.

The primary cause of gas meter failure was the selection of the gas meter type. The gas meter installed at the site was a rotary displacement gas meter which is not recommended by 10 State Standards (2003) for use with biogas. The tight tolerances of the rotary displacement type gas meter results in rotor binding due to gas particulates and excess torque placed on the meter housing from installation. Inspection of the gas meter revealed small particles that were being carried by the biogas would become lodged in the rotor causing the rotor to stall. Secondly, the design of the aluminum rotary gas meter requires that the meter be securely mounted and inlet and outlet piping must be separately supported. The initial installation of gas meter piping was supported by the meter itself resulting in binding of the internal rotor. Flexible piping and additional piping supports were installed. However, due to space restrictions piping could not be completely supported.

Due to the moist nature of the biogas, it is recommended that gas handling equipment be protected from the environment. This will aid in the controlling the location of water vapor condensation within the system and prevent freezing of the condensate. Additionally, the selection of a gas metering components should be rated for use with biogas and installed per manufacturer's specifications.

Reactor Cover

There are three options for the design of the reactor cover and biogas gas storage component; flexible membrane cover, fixed roof with external biogas storage, and floating cover. The primary design criteria requirement is the ability for the component(s) to withstand a pressure and volume change equivalent to the cycle treatment depth and volume. For example a 1,000 m^3 reactor with a diameter of 12.2 m and operating depth of 8.5 meters operating at two cycles per day will produce a pressure change of 0.86 m W.C. (33.9” W.C.) and volume change of 100 m^3 during the decant and fill phases, thus requiring a fixed cover reactor to have external biogas storage to regulate the pressure change within the reactor.

The OSU SREC ASBR and ISU ASBR were designed and constructed with a flexible membrane cover. As the ASBR is a batch process reactor, at the end of each cycle the cycle treatment volume is removed prior to the addition of fresh influent. The advantages of the flexible membrane cover are the simplicity of installation and cost. Although less expensive than fixed and floating covers; the expected design life of a flexible membrane cover is much shorter. The reduction in operational life of the cover is due to continuous stresses from the expansion and contractions of the cover during decant and feed phases. These stresses were observed at the OSU SREC ASBR by the tearing of the membrane along the perimeter of the cover where the mounting hardware was located. These tears were observed within the first year of operation and within 7 years of installation.

Based upon the operational experience at the OSU SREC ASBR, the flexible membrane cover is not recommended for installation on an ASBR. A solid fixed cover with external biogas storage or floating cover is recommended due to life expectancy concern of the flexible membrane cover. Additionally, the puncturing of the flexible membrane cover also poses a potential safety concern due release of biogas, creating a potential fire and explosion hazard.

Settling

The settling phase length is a function of the zone settling velocity and settling depth. The settling distance is the distance the zone settling front must travel to reach the desired level within the reactor prior to release of effluent, during the decant phase. The required settling depth is a function of the reactor geometry, HRT, and cycle length (T_{cycle}). The calculation steps for determining the settling distance (H_{sd}, m) are the given in eq. 33, 34, and 35 of and illustrated in figure 39.

$$H_{sd} = (t_c\ H_{ro}) / HRT \tag{33}$$

Where:

H_{ro} = Reactor operating height, m

$$H_{ro} = V_{ro} \text{ x } 4 / \pi\ / D^2 \tag{34}$$

Where:

D= Reactor diameter, m

$$H_{sd} = V_c \text{ x } 4 / \pi / D^2 \tag{35}$$

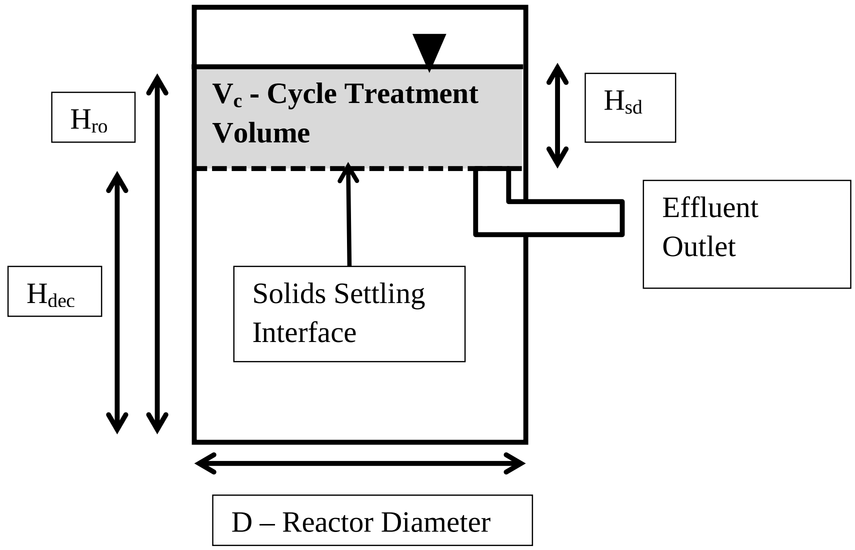

Figure 39. Settling depth of an ASBR

Given a known reactor geometry, HRT, cycle length and settling velocity, the minimum settling phase length (t_{smin}) is a determined (eq. 36). As discussed previously, the settling velocity is not constant but a function of the substrate and mixed liquor solids concentration. The height of the hindered solids settling interface with respect to time and reactor operating height is shown in figure 40. As the solids concentration increases in the reactor, settling velocity will decrease and the height and time that the transition and compression settling occurs will increase and decrease respectively. Thus a maximum settling time (t_{smax}) is determined based upon the time to reach transition settling (eq. 37)

$$t_{smin} = H_{sd} / v_s \tag{36}$$

Where:

v_s – Solids settling velocity, m/min

$$t_{smax} = t_t - t_d \quad (37)$$

where:

t_t – Time required to reach transition settling

t_{comp} – Time required to reach compression settling

t_d – Decant phase length

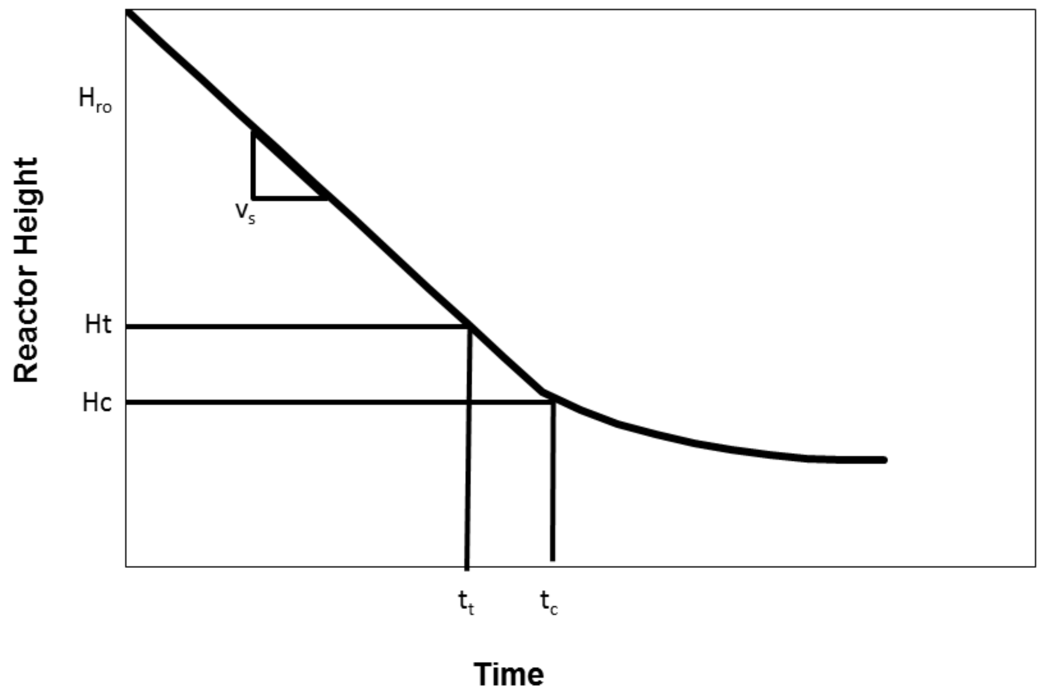

Figure 40. Zone solids settling solids interface height with respect to time

As the operational performance of the ASBR is dependent upon internal solids separation, the settling velocity and settling depth will determine the reactor operating height for a given influent and design mixed liquor solids concentration. Given a solids settling curve as shown in figure 38, and t_{smax}, the maximum reactor operating height H_{romax} can be determined with respect to cycle time and HRT (eq. 38).

$$H_{romax} = t_{smax} \times v_s \times HRT / t_c \qquad (38)$$

A reactor with a mixed liquor settling velocity of 0.02 m/min, a t_{smax} of 90 and operating at a 5 day HRT at two cycles per day has a H_{romax} of 18 m (59 ft). This however may not be practical due to reactor height limitations for mixing. Reactor dimension limitations presented by Bathija (1982) for basic jet mixing design limits reactor operating heights to 9.1 m (30 ft) and diameters to 12.2 m (40 ft), thus limiting total reactor volumes to 1,360 m^3. Utilizing these reactor dimension limitations allows for settling depths of 1.8 and 0.9 m at 1 and 2 cycles per day at 5 day HRT. Given the limitation of the reactor size due to mixing system limitations the use multiple reactors may be required.

Sludge Handling

As the mixed liquor solids reach the design concentration, wasting of settled concentrated solids is necessary to maintain the mixed liquor solids concentrations within the optimum range. Maintaining solids concentrations within this range ensures consistent settling and biological activity. The excess solids are removed from the bottom of the reactor after the completion of the settling phase. The sludge wasting volume for removal is estimated based upon the solids concentration of the settled portion of the mixed liquor. After removal of the prescribed sludge volume, the remaining cycle volume is removed from the effluent withdrawal port, thus completing the sludge removal and decant phase. The time period between sludge wasting is based upon the sludge production rate and the concentration of the settled solids that can be removed from the bottom of the reactor via the sludge removal port. Determination of the sludge wasting period t_s (days) is as follows:

$$t_s = (V_{sludge}) \times (SMLS / SPR) \quad (39)$$

Where:

V_{sludge} – Volume of sludge removed (settled mixed liquor solids), m^3

SMLS – Settled mixed liquor solids concentration, kg m^{-3}

SPR – Sludge production rate, kg day^{-1}

The settled mixed liqour solids concentration can be estimated as follows:

$$SMLS = (H_{ro} \times MLS - H_S \times C_{eff}) / H_D \quad (40)$$

Where:

H_{ro} – Mixed liquor react phase height, m

H_{dec} – Mixed liqour decant phase height, m

C_{eff} – Effluent solids concentration, kg m^{-3}

MLS – Mixed liquor solids concentration, kg m^{-3}

SMLS – Settled mixed liquor solids concentration, kg m^{-3}

Each reactor and influent combination will result in a unique maximum mixed liquor solids concentration. The maximum mixed liquor solids concentration is the solids concentration obtained at which settling is hindered, resulting in solids washout. The utilization of scheduled sludge wasting provides the ability to maintain an average mixed liquor solids concentration slightly below the maximum ensuring consistent solids settling and retention.

Reactor Serviceability

There are two management strategies for the operation of the completed ASBR system, contracted service and in-house management. The management decision for the system operation must be considered during the design process as specification choices for some system components will favor one management style over the other. For in-house management, system components specifications should match those of the existing facility and expertise of local professionals. The ability to utilize local service professionals and component suppliers aids in reducing maintenance down times.

Conclusions

The design of an effective ASBR centers on two components; energy balance and solids settling. The energy inputs for temperature maintenance, mixing, and influent and effluent transfer must be equal to or less than the potential recoverable energy. Influent waste streams with inadequate recoverable energy content to facilitate ASBR operation results in a negative energy balance. Thus eliminating the potential for excess energy generation for utilization by the facility or sale to local utility provider

The internal settling and retention of solids in the ASBR are the key parameters in the physical design and operation of the ASBR. Internal solids retention and resulting separation of the HRT and SRT distinguishes the ASBR from other anaerobic digestion reactors. As the solids settling velocity is a function of the mixed liquor solids concentration, it will determine design operating mixed liquor solids concentration. The mixing system power requirements and design are a function of the mixed liquor solids concentration and solids settling velocity and also set limitations to the reactor geometry. The solids settling velocity of the mixed liquor solids within the reactor are the defining characteristic of operational performance of the ASBR and its process design. The solids retention of the ASBR provides the option for flexible operating temperature selection. Influent heating at temperatures less than 35˚C can provide the ASBR the potential for use of low strength influent waste streams.

Utilizing solids settling and the reactor energy balance as the basic universal design principles, an individual ASBR design can be developed. Inclusion of both design principles will provide adequate data for the determination of the ASBR applicability for individual waste streams and facility locations.

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APPENDICES

Appendix 1. OSU SREC ASBR biogas production data

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
8/19/08 12:10 PM	42.81	
8/20/08 10:30 AM	70.00	2,922
8/20/08 4:20 PM	73.74	1,539
8/21/08 9:30 AM	73.74	
8/21/08 3:50 PM	82.66	3,380
8/22/08 8:25 AM	99.80	2,481
8/22/08 1:00 PM	104.99	2,718
8/23/08 9:00 AM	127.18	2,663
8/24/08 12:00 AM		
8/25/08 8:00 AM	128.02	
8/26/08 8:25 AM	128.02	
8/26/08 11:50 AM	129.63	
8/27/08 8:15 AM	148.81	2,255
8/28/08 9:30 AM	174.74	2,465
8/29/08 8:00 AM	191.62	1,801
9/2/08 10:10 AM	328.20	3,339
9/3/08 8:00 AM	353.63	2,795
9/4/08 9:30 AM	382.87	2,752
9/5/08 8:00 AM	405.31	2,394
9/8/08 8:50 AM	497.25	3,030
9/9/08 9:05 AM	532.99	3,537
9/10/08 8:55 AM	559.23	2,642
9/11/08 9:30 AM	585.59	2,573
9/11/08 4:45 PM	593.32	2,559
9/12/08 9:15 AM	609.90	2,412
9/15/08 4:15 PM	738.06	3,893
9/16/08 8:15 AM	759.46	3,210

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
9/17/08 10:05 AM	760.18	
9/23/08 9:45 AM	783.68	
9/26/08 8:20 AM	783.72	
9/29/08 9:10 AM	893.84	3,629
9/30/08 8:05 AM	925.67	3,333
10/1/08 4:55 PM	976.32	3,702
10/2/08 8:00 AM	992.67	2,602
10/3/08 8:15 AM	1,030.92	3,786
10/6/08 9:30 AM	1,151.41	3,948
10/7/08 8:10 AM	1,186.07	3,670
10/8/08 8:20 AM	1,215.83	2,955
10/9/08 8:35 AM	1,258.98	4,271
10/10/08 9:35 AM	1,307.98	4,704
10/13/08 8:35 AM	1,417.69	3,709
10/14/08 4:25 PM	1,425.56	593
10/15/08 11:45 AM	1,445.45	2,469
10/16/08 8:20 AM	1,471.57	3,046
10/17/08 2:30 PM	1,501.44	2,376
10/20/08 8:20 AM	1,598.68	3,545
10/21/08 4:15 PM	1,641.77	3,240
10/22/08 8:10 AM	1,664.54	3,433
10/23/08 9:25 AM	1,692.60	2,667
10/24/08 12:45 PM	1,716.53	2,101
10/27/08 8:30 AM	1,765.91	1,749
10/28/08 2:40 PM	1,772.59	531
10/29/08 8:05 AM	1,783.11	1,450
10/30/08 8:20 AM	1,805.09	2,175
10/31/08 8:35 AM	1,823.06	1,778
11/3/08 9:10 AM	1,893.41	2,326
11/4/08 10:35 AM	1,918.98	2,414
11/5/08 9:00 AM	1,946.14	2,908
11/6/08 9:55 AM	1,978.29	3,097
11/7/08 8:00 AM	1,993.87	1,693
11/10/08 8:20 AM	2,052.28	1,938
11/11/08 8:30 AM	2,075.41	2,297

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
11/12/08 8:55 AM	2,100.63	2,479
11/13/08 8:10 AM	2,123.12	2,322
11/14/08 8:25 AM	2,150.94	2,753
11/17/08 8:05 AM	2,208.39	1,924
11/18/08 5:20 PM	2,239.96	2,279
11/19/08 8:30 AM	2,249.96	1,582
11/20/08 8:50 AM	2,279.93	2,956
11/21/08 4:40 PM	2,299.20	1,453
11/25/08 12:30 PM	2,413.14	2,978
11/26/08 5:50 PM	2,452.71	3,238
12/1/08 8:20 AM	2,627.10	3,788
12/2/08 8:25 AM	2,668.51	4,127
12/3/08 8:10 AM	2,712.67	4,462
12/4/08 8:35 AM	2,731.92	1,892
12/8/08 11:20 AM	2,861.71	3,154
12/9/08 2:15 PM	2,900.17	3,429
2/2/09 3:40 PM	3,100.01	
2/3/09 10:30 AM	3,109.72	1,237
2/4/09 1:10 PM	3,115.09	483
2/5/09 9:00 AM	3,115.09	0
2/6/09 3:25 PM	3,115.09	0
2/7/09 11:50 AM	3,139.60	2,881
2/9/09 8:35 AM	3,196.28	3,040
2/10/09 9:30 AM	3,227.40	2,998
2/11/09 8:10 AM	3,255.40	2,965
2/12/09 8:20 AM	3,278.39	2,283
2/17/09 9:45 AM	3,438.39	3,163
2/18/09 9:05 AM	3,467.58	3,002
2/19/09 8:35 AM	3,489.60	2,249
2/20/09 11:35 AM	3,520.47	2,744
2/23/09 9:05 AM	3,558.65	1,318
2/24/09 8:25 AM	3,583.17	2,522
2/25/09 9:40 AM	3,607.86	2,347
2/26/09 9:40 AM	3,633.09	2,523
2/27/09 8:45 AM	3,645.82	1,324
3/2/09 8:30 AM	3,675.59	996

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
3/3/09 9:00 AM	3,695.66	1,966
3/4/09 8:45 AM	3,721.33	2,594
3/5/09 8:20 AM	3,735.57	1,449
3/6/09 8:50 AM	3,761.93	2,582
3/9/09 8:30 AM	3,828.80	2,239
3/10/09 8:15 AM	3,834.68	594
3/11/09 8:40 AM	3,854.22	1,921
3/12/09 9:20 AM	3,855.07	83
3/13/09 10:15 AM	3,879.65	2,368
3/17/09 8:25 AM	3,928.32	1,240
3/18/09 9:25 AM	3,944.75	1,577
3/19/09 9:00 AM	3,946.00	127
3/20/09 8:55 AM	3,969.01	2,309
3/23/09 9:10 AM	4,033.75	2,151
3/24/09 8:55 AM	4,045.77	1,215
3/25/09 9:05 AM	4,049.28	349
3/30/09 8:45 AM	4,074.90	514
3/31/09 8:40 AM	4,087.49	1,263
4/1/09 8:15 AM	4,101.06	1,381
4/2/09 8:40 AM	4,120.28	1,889
4/3/09 8:25 AM	4,133.01	1,286
4/7/09 8:15 AM	4,149.40	410
4/9/09 8:45 AM	4,152.20	139
4/13/09 11:15 AM	4,181.27	708
4/14/09 11:30 AM	4,187.16	583
4/15/09 10:15 AM	4,191.24	430
4/16/09 10:55 AM	4,195.23	388
4/17/09 11:30 AM	4,201.65	627
4/20/09 9:00 AM	4,219.41	613
4/22/09 10:00 AM	4,219.41	0
4/23/09 8:45 AM	4,247.73	2,988
4/24/09 8:40 AM	4,279.03	3,141
4/28/09 8:05 AM	4,409.86	3,291
4/29/09 12:10 PM	4,444.27	2,941
4/30/09 8:20 AM	4,472.99	3,418
5/1/09 8:25 AM	4,500.83	2,774

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
5/4/09 8:45 AM	4,585.86	2,821
5/6/09 8:30 AM	4,639.56	2,699
5/7/09 8:15 AM	4,657.79	1,842
5/8/09 6:40 AM	4,688.89	3,330
5/12/09 8:30 AM	4,817.98	3,167
5/13/09 8:15 AM	4,857.40	3,983
5/14/09 8:40 AM	4,892.05	3,406
5/15/09 8:55 AM	4,926.45	3,405
5/18/09 11:41 AM	5,021.23	3,042
5/19/09 8:05 AM	5,043.25	2,591
5/20/09 9:00 AM	5,067.77	2,362
5/21/09 8:15 AM	5,091.28	2,427
5/22/09 7:50 AM	5,116.63	2,580
5/26/09 9:00 AM	5,202.02	2,109
5/27/09 8:35 AM	5,228.92	2,738
5/28/09 8:55 AM	5,260.81	3,145
5/29/09 8:00 AM	5,303.47	4,435
6/1/09 12:00 AM	5,441.51	5,177
6/2/09 8:50 AM	5,463.88	1,635
6/3/09 8:30 AM	5,499.07	3,569
6/4/09 8:15 AM	5,542.38	4,377
6/5/09 8:05 AM	5,591.95	4,992
6/8/09 7:55 AM	5,680.70	2,965
6/9/09 8:00 AM	5,711.70	3,089
6/10/09 8:05 AM	5,749.11	3,728
6/11/09 8:30 AM	5,809.69	5,955
6/12/09 8:15 AM	5,855.63	4,642
6/15/09 8:20 AM	5,875.66	667
6/17/09 11:30 AM	5,945.31	3,267
6/18/09 8:15 AM	5,976.36	3,591
6/19/09 11:00 AM	6,028.34	4,664
6/25/09 10:00 AM	6,117.67	1,499
6/29/09 7:45 AM	6,117.67	0
6/30/09 8:10 AM	6,119.34	
7/1/09 8:00 AM	6,135.83	
7/2/09 8:30 AM	6,145.42	

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
7/7/09 8:45 AM	6,222.71	1,543
7/7/09 2:20 PM	6,237.80	
7/7/09 2:20 PM	0.43	
7/9/09 8:10 AM	34.48	2,487
7/10/09 8:50 AM	97.48	6,130
7/16/09 9:00 AM	309.37	3,527
7/17/09 8:10 AM	344.56	3,646
7/20/09 8:15 AM	406.31	2,056
7/21/09 8:35 AM	420.14	1,364
7/22/09 8:10 AM	448.86	2,923
7/23/09 8:30 AM	496.03	4,652
7/24/09 8:30 AM	546.61	5,058
7/27/09 8:50 AM	629.41	2,747
7/29/09 8:30 AM	653.41	1,208
7/30/09 9:10 AM	671.68	1,778
7/31/09 11:55 AM	705.73	3,055
8/3/09 8:30 AM	770.22	2,257
8/4/09 8:10 AM	795.80	2,594
8/5/09 8:30 AM	828.59	3,234
8/7/09 8:30 AM	865.41	1,841
8/10/09 8:30 AM	989.30	4,130
8/11/09 8:45 AM	1,019.96	3,034
8/12/09 1:50 PM	1,061.65	3,440
8/17/09 8:30 AM	1,224.69	3,412
8/19/09 8:45 AM	1,251.65	1,341
8/20/09 8:30 AM	1,277.45	2,607
8/21/09 8:35 AM	1,291.36	1,386
8/24/09 9:10 AM	1,359.95	2,268
8/25/09 8:00 AM	1,377.76	1,872
8/26/09 8:20 AM	1,382.33	451
8/28/09 11:30 AM	1,383.83	70
8/31/09 8:15 AM	1,462.88	2,760
9/1/09 3:15 PM	1,484.10	1,643
9/2/09 8:40 AM	1,486.82	375
9/4/09 3:45 PM	1,571.46	3,688
9/8/09 9:00 AM	1,682.64	2,990

Appendix 1. OSU SREC ASBR biogas production data continued

Date	Gas meter Reading	Daily Biogas Production
	x 100 ft^3	ft^3/day
9/11/09 12:00 AM	1,759.61	2,932
9/14/09 1:50 PM	1,857.98	2,751
9/15/09 3:55 PM	1,857.98	0
9/16/09 9:00 AM	1,857.98	0
9/18/09 2:25 PM	1,857.98	0
9/21/09 1:00 PM	1,936.64	2,675

Appendix 2. OSU SREC ASBR Influent parameters, operating levels, and volumes

Date	HRT	React Phase Depth	Decant Depth	Reactor Working Volume	Influent Volume	TS	TVS	COD	PH	OLR	OLR
	days	ft	ft	L	L	mg/l	mg/l	mg/l		g VS/l/d	g COD/l/d
2/3/2009	20	11.40	0.57	405,726	20,286	14,374	10,916	19,691	6.57	0.55	0.98
2/5/2009	20	11.40	0.57	405,726	20,286	14,694	11,151	21,422	6.53	0.56	1.07
2/20/2009	20	11.40	0.57	405,726	20,286	14,054	10,750	18,693	6.52	0.54	0.93
2/23/2009	20	11.40	0.57	405,726	20,286		3,264	4,763	7.29	0.16	0.24
2/24/2009	20	11.40	0.57	405,726	20,286	5,698	4,022	6,457	7.32	0.20	0.32
2/27/2009	20	11.40	0.57	405,726	20,286	8,218	5,752	11,728	6.88	0.29	0.59
3/4/2009	20	11.40	0.57	405,726	20,286	12,168	9,009	16,283	6.62	0.45	0.81
3/6/2009	20	11.40	0.57	405,726	20,286	10,276	7,054	13,874	6.55	0.35	0.69
3/12/2009	20	11.40	0.57	405,726	20,286						
3/17/2009	20	11.40	0.57	405,726	20,286	6,159	3,819	7,153	6.83	0.19	0.36
3/24/2009	20	11.40	0.57	405,726	20,286	6,210	4,298	4,800	7.14	0.21	0.24
3/31/2009	20	11.40	0.57	405,726	20,286	16,414	13,386	14,834	7.13	0.67	0.74
4/7/2009	20	11.40	0.57	405,726	20,286	5,178	3,333	5,534	7.19	0.17	0.28
4/13/2009	20	11.40	0.57	405,726	20,286	8,181	5,576	13,610	6.69	0.28	0.68
4/17/2009	20	11.40	0.57	405,726	20,286	11,430	7,937	18,693	6.64	0.40	0.93
4/22/2009	20	11.40	0.57	405,726	20,286	9,621	7,746	14,664	6.84	0.39	0.73
5/1/2009	20	11.40	0.57	405,726	20,286			9,600	6.88		0.48
5/6/2009	20	11.40	0.57	405,726	20,286	12,061	9,379	18,523	6.72	0.47	0.93
5/15/2009	20	11.40	0.57	405,726	20,286	8,429	5,987	13,893	6.70	0.30	0.69
5/21/2009	20	11.40	0.57	405,726	20,286	8,148	5,598	10,918	7.22	0.28	0.55
5/28/2009	20	11.40	0.57	405,726	20,286	14,361	10,316	20,444	6.89	0.52	1.02
6/4/2009	20	11.40	0.57	405,726	20,286	9,874	6,909	15,267	6.68	0.35	0.76

Appendix 2. OSU SREC ASBR Influent parameters, operating levels, and volumes continued

Date	HRT	React Phase Depth	Decant Depth	Reactor Working Volume	Influent Volume	TS	TVS	COD	PH	OLR	OLR
	days	ft	ft	L	L	mg/l	mg/l	mg/l		g VS/l/d	g COD/l/d
6/11/2009	20	11.40	0.57	405,726	20,286	9,978	7,241	15,926	6.85	0.36	0.80
6/18/2009	20	11.40	0.57	405,726	20,286	8,801	6,098	14,344	6.63	0.30	0.72
7/1/2009	20	11.40	0.57	405,726	20,286	6,680	4,200			0.21	
7/9/2009	20	11.40	0.57	405,726	20,286	9,433	6,794			0.34	
7/16/2009	20	11.40	0.57	405,726	20,286	5,570	3,820		7.07	0.19	
7/23/2009	19	10.70	0.56	380,813	20,043	9,042	6,168		6.82	0.32	
7/30/2009	18	10.10	0.56	359,459	19,970	8,849	6,084		6.77	0.34	
8/5/2009	17	9.50	0.56	338,105	19,889	7,940	5,263		7.00	0.31	
8/10/2009	16	8.91	0.56	317,107	19,819					0.00	
8/26/2009	15	8.93	0.60	317,819	21,188	7,040	4,881		6.73	0.33	
9/2/2009	15	8.93	0.60	317,819	21,188	11,540	8,023		6.70	0.53	
9/11/2009	15	8.93	0.60	317,819	21,188	8,509	5,846	14,533		0.39	0.97
9/16/2009	14	8.95	0.64	318,531	22,752	11,571	8,597	15,681	6.67	0.61	
9/22/2009	13	8.98	0.69	319,598	24,584	5,663	3,481		7.20	0.27	
9/30/2009	12	9.01	0.75	320,666	26,722	9,230	6,141	11,370	7.13	0.51	0.95
10/7/2009	11	9.04	0.82	321,734	29,249	12,749	9,439	17,300	6.84	0.86	1.57
10/14/2009	10	9.08	0.91	323,157	32,316	10,092	7,251	12,820	6.81	0.73	1.28
10/23/2009	10	9.08	0.91	323,157	32,316	11,962	8,667	18,956	6.55	0.87	1.90
10/28/2009	10	9.08	0.91	323,157	32,316	7,506	5,470			0.55	
11/6/2009	10	9.08	0.91	323,157	32,316	6,521	4,479			0.45	
11/13/2009	9	9.14	1.02	325,293	36,144	4,006	2,510			0.28	
12/2/2009	6	9.41	1.57	334,902	55,817	4,006	2,510			0.42	
12/16/2009	5	9.59	1.92	341,308	68,262	3,173	1,820		7.21	0.36	
1/25/2010	5	9.59	1.92	341,308	68,262	3,537	2,192	6,494	7.33	0.44	1.30
4/22/2010	5	9.59	1.92	341,308	68,262	4,859	3,159			0.63	

Appendix 3. OSU SREC Mixed Liquor Parameters

Date	TS mg/l	TVS mg/l	COD mg/l	PH	TS kg	TVS kg	COD kg
1/23/2009	5,848	3,896			2,372	1,580	
1/26/2009	6,491	4,317			2,633	1,751	
2/2/2009	4,717	2,916	6,165	6.97	1,914	1,183	2,501
2/5/2009	5,123	3,174	8,189	6.97	2,079	1,288	3,322
2/10/2009	6,448	4,242	7,163	6.97	2,616	1,721	2,906
2/12/2009	5,894	3,894	6,824	6.99	2,391	1,580	2,769
2/17/2009	6,221	4,234	7,238	6.99	2,524	1,718	2,937
2/20/2009	5,777	3,690	7,482	6.92	2,344	1,497	3,036
2/23/2009	6,737	4,482	8,057	6.94	2,733	1,818	3,269
2/24/2009	6,384	4,252	7,247	6.94	2,590	1,725	2,940
2/27/2009	5,693	3,886	6,099	7.02	2,310	1,577	2,475
3/3/2009	6,383	4,382	6,335	6.96	2,590	1,778	2,570
3/4/2009	6,149	4,343	7,210	7.01	2,495	1,762	2,925
3/5/2009	6,512	4,187	11,050	6.90	2,642	1,699	4,483
3/6/2009	6,541	4,351	5,224	6.90	2,654	1,765	2,120
3/9/2009	6,668	4,533	7,497	6.96	2,705	1,839	3,042
3/11/2009	6,877	4,494	7,215	6.96	2,790	1,823	2,927
3/12/2009	6,399	4,197	7,017	7.00	2,596	1,703	2,847
3/17/2009	6,368	4,280	7,102	6.95	2,584	1,737	2,881
3/18/2009	7,276	4,935		7.00	2,952	2,002	
3/19/2009	7,204	4,876	5,605	6.97	2,923	1,978	2,274
3/20/2009	6,127	3,958			2,486	1,606	
3/23/2009	6,041	3,873		6.98	2,451	1,571	
3/24/2009	6,054	3,758	6,353	6.99	2,456	1,525	2,578
3/25/2009	5,724	3,534		6.99	2,322	1,434	
3/26/2009	5,888	3,713	5,407	6.96	2,389	1,506	2,194
3/31/2009	6,046	3,993	6,170	7.04	2,453	1,620	2,503
4/7/2009	5,172	3,312	4,843	7.02	2,098	1,344	1,965
4/9/2009	5,072	3,226		7.01	2,058	1,309	
4/17/2009	10,116	7,647	11,365	6.90	4,104	3,103	4,611
4/20/2009	9,366	7,416			3,800	3,009	
4/22/2009	6,171	4,403	6,306	6.93	2,504	1,786	2,559
4/24/2009	8,106	6,092	10,787		3,289	2,472	4,377
5/1/2009			5,647	7.11			2,291
5/6/2009	5,846	4,061	6,419	6.95	2,372	1,648	2,604
5/15/2009	5,893	3,994	6,156	7.02	2,391	1,620	2,498
5/21/2009	5,154	3,248	5,120	7.13	2,091	1,318	2,077

Appendix 3. OSU SREC Mixed Liquor Parameters continued

	TS	TVS	COD	PH	TS	TVS	COD
Date	mg/l	mg/l	mg/l		kg	kg	kg
5/28/2009	5,873	3,980	5,798	7.10	2,383	1,615	2,352
6/4/2009	5,911	3,740	5,949	7.04	2,398	1,517	2,414
6/11/2009	6,644	4,528	6,815	7.13	2,696	1,837	2,765
6/18/2009	6,482	4,354	6,419	7.05	2,630	1,767	2,604
7/1/2009	6,944	4,540			2,817	1,842	
7/9/2009	7,423	4,944			3,012	2,006	
7/16/2009	7,838	5,282		7.16	3,180	2,143	
7/23/2009	9,482	6,586		7.05	3,611	2,508	
7/30/2009	12,012	8,734		6.99	4,318	3,140	
8/5/2009	8,789	5,948		7.09	2,972	2,011	
8/10/2009	9,922				3,146		
8/19/2009	7,971	5,312			2,533	1,688	
8/26/2009	9,283	6,499			2,950	2,066	
9/2/2009	12,684	9,204		6.98	4,031	2,925	
9/11/2009	9,526	6,398	12,274		3,028	2,033	3,901
9/16/2009	9,119	6,350	9,601	7.08	2,905	2,023	3,058
9/22/2009	10,081	6,960	9,883	7.19	3,222	2,224	3,159
9/30/2009	10,570	7,363	11,408	7.13	3,389	2,361	3,658
10/7/2009	10,590	7,324	11,314	6.98	3,407	2,356	3,640
10/14/2009	9,943	7,172	10,749	6.93	3,213	2,318	3,474
10/23/2009	9,621	7,026	11,200	6.92	3,109	2,271	3,619
10/28/2009	10,178	7,488			3,289	2,420	
11/6/2009	9,371	6,771			3,028	2,188	
11/13/2009	9,456	6,331			3,076	2,059	
12/2/2009	11,646	8,816			3,900	2,952	
12/16/2009	8,549	6,090		7.05	2,918	2,079	
1/25/2010	6,347	4,419		6.92	2,166	1,508	
4/6/2010	6,683	4,380			2,281	1,495	
4/13/2010	6,274	4,149			2,141	1,416	
4/22/2010	6,896	4,693			2,354	1,602	

Appendix 4. OSU SREC ASBR Effluent Parameters

Date	TS mg/l	TVS mg/l	COD mg/l	PH
1/26/2009	4,843	2,909		
2/5/2009	4,169	2,232	3,680	7.04
2/10/2009	5,040	3,040	5,553	6.99
2/12/2009	4,666	2,763	5,572	7.03
2/17/2009	5,244	3,293	6,645	7.00
2/20/2009	5,228	3,286	5,854	6.99
2/24/2009	5,129	3,151	6,193	7.03
2/27/2009	4,226	2,534	4,386	7.09
3/3/2009	5,008	3,121	3,737	7.03
3/4/2009	5,056	3,117	5,506	6.97
3/6/2009	5,352	3,274	4,919	6.98
3/9/2009	5,108	3,137	5,064	6.96
3/11/2009	5,473	3,340	5,563	7.00
3/12/2009	4,962	2,937	4,951	7.10
3/18/2009	5,577	3,433		7.05
3/19/2009	4,918	2,847	4,330	7.09
3/20/2009	5,009	2,966		
3/23/2009	4,158	2,318		7.13
3/25/2009	3,399	1,558		7.07
3/26/2009	4,222	2,411	3,360	7.10
3/27/2009	3,817	1,948		
3/31/2009	4,941	3,019	4,989	7.14
4/7/2009	2,857	1,317	1,807	7.11
4/17/2009	5,587	3,748	6,419	7.00
4/24/2009	4,737	2,964	5,393	
5/1/2009			3,897	7.17
5/6/2009	4,650	2,954	4,744	7.05
5/15/2009	4,407	2,650	4,334	7.15
5/21/2009	3,341	1,891	2,838	7.17
5/28/2009	4,507	2,778	3,741	7.20
6/4/2009	3,974	2,220	3,374	7.17
6/11/2009	5,504	3,498	5,210	7.25
6/18/2009	5,732	3,580	5,732	7.17
7/1/2009	2,722	1,286		
7/9/2009	2,861	1,292		
7/16/2009	2,974	1,384		7.33
7/23/2009	2,594	1,129		7.16

Appendix 4. OSU SREC ASBR Effluent Parameters continued

	TS	TVS	COD	PH
Date	mg/l	mg/l	mg/l	
7/30/2009	2,656	1,123		7.10
8/5/2009	7,343	4,720		7.17
8/19/2009	3,248	1,604		
8/26/2009	2,259	1,146		
9/2/2009	2,057	777		7.05
9/11/2009	6,437	3,854	6,532	
9/22/2009	2,853	1,152	1,901	7.25
9/30/2009	3,238	1,508	2,979	7.26
10/7/2009	7,599	4,758	7,835	7.11
10/14/2009	5,270	3,153		
10/28/2009	3,712	2,062		
11/6/2009	4,688	2,761		
12/2/2009	2,551	1,031		
12/16/2009	2,184	871		7.23
1/25/2010	4,797	3,141		7.11
4/22/2010	2,950	1,274		

Appendix 5. OSU SREC ASBR organic matter mass balance data

Date	Actual Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
2/2/2009		0.00	65	0.00	59.01	0.00	0.00	1183.00	22.17
2/3/2009	82.00	82.00	65	221.45	59.01	89.58	0.00	1255.85	
2/4/2009	82.00	82.00	65	226.21	59.01	89.58	62.79	1270.68	
2/5/2009	82.00	82.00	65	226.21	45.28	89.58	63.53	1298.50	27.64
2/6/2009	82.00	82.00	65	226.21	48.55	89.58	0.00	1386.59	
2/7/2009	81.59	81.59	65	226.21	51.81	89.13	0.00	1471.86	
2/8/2009	82.00	82.00	65	226.21	55.08	89.58	0.00	1553.42	
2/9/2009	86.08	86.08	65	226.21	58.34	94.03	0.00	1627.25	
2/10/2009	84.88	84.88	65	226.21	61.67	92.73	0.00	1699.07	33.76
2/11/2009	83.95	83.95	65	226.21	58.85	91.71	0.00	1774.72	
2/12/2009	64.65	64.65	65	226.21	56.05	70.63	0.00	1874.25	34.14
2/13/2009	84.00	84.00	65	226.21	58.20	91.76	0.00	1950.50	
2/14/2009	84.00	84.00	65	226.21	60.35	91.76	0.00	2024.60	
2/15/2009	84.00	84.00	65	226.21	62.50	91.76	0.00	2096.54	
2/16/2009	84.00	84.00	65	226.21	64.65	91.76	0.00	2166.34	
2/17/2009	89.56	89.56	65	226.21	66.80	97.83	0.00	2227.91	34.33
2/18/2009	85.02	0.00	65	0.00	66.80	0.00	0.00	2161.11	
2/19/2009	63.68	0.00	65	0.00	66.80	0.00	0.00	2094.31	
2/20/2009	77.70	77.70	65	218.08	66.66	84.88	0.00	2160.84	30.65
2/21/2009	0.00	0.00	65	0.00	66.66	0.00	0.00	2094.18	
2/22/2009	0.00	0.00	65	104.18	66.66	0.00	0.00	2131.70	
2/23/2009	37.33	37.33	65	66.21	66.66	40.79	0.00	2090.47	37.93
2/24/2009	71.42	71.42	65	81.59	63.92	78.02	0.00	2030.12	37.45
2/25/2009	66.45	66.45	65	99.14	59.76	72.60	0.00	1996.90	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
2/26/2009	71.44	71.44	65	99.14	55.60	78.05	0.00	1962.39	
2/27/2009	66.00	66.00	65	116.69	51.41	72.10	0.00	1955.57	41.49
2/28/2009	66.00	66.00	65	129.87	54.45	72.10	0.00	1958.89	
3/1/2009	66.00	66.00	65	143.06	57.49	72.10	0.00	1972.36	
3/2/2009	66.00	66.00	65	156.25	60.53	72.10	0.00	1995.97	
3/3/2009	55.67	55.67	65	169.43	63.31	60.82	0.00	2041.27	40.20
3/4/2009	73.45	73.45	65	182.76	63.23	80.24	0.00	2080.55	40.04
3/5/2009	41.04	41.04	65	162.92	63.23	44.83	0.00	2135.41	38.74
3/6/2009	73.12	73.12	65	143.10	66.42	79.88	0.00	2132.21	38.74
3/7/2009	63.00	63.00	65	133.77	66.42	68.82	0.00	2130.74	
3/8/2009	63.00	63.00	65	124.44	66.42	68.82	0.00	2119.93	
3/9/2009	63.41	63.41	65	115.10	63.64	69.27	0.00	2102.13	42.27
3/10/2009	63.00	63.00	65	105.77	63.64	68.82	0.00	2075.44	
3/11/2009	54.39	54.39	65	96.44	67.76	59.41	0.00	2044.71	39.92
3/12/2009	54.00	54.00	65	87.11	59.58	58.99	0.00	2013.24	41.79
3/13/2009	67.04	67.04	65	77.47	61.26	73.24	0.00	1956.21	
3/14/2009	67.00	67.00	65	77.47	62.95	73.19	0.00	1897.54	
3/15/2009	67.00	67.00	65	77.47	64.63	73.19	0.00	1837.19	
3/16/2009	67.00	67.00	65	77.47	66.32	73.19	0.00	1775.16	
3/17/2009	35.13	35.13	65	77.47	68.00	38.37	88.76	1657.50	37.64
3/18/2009	44.66	44.66	65	77.47	69.64	48.79	0.00	1616.54	42.61
3/19/2009	65.00	65.00	65	77.47	57.76	71.01	0.00	1565.25	49.59
3/20/2009	65.38	65.38	65	77.47	60.17	71.43	0.00	1511.13	38.97
3/21/2009	61.00	61.00	65	77.47	54.08	66.64	0.00	1467.88	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
3/22/2009	61.00	61.00	65	77.47	48.00	66.64	0.00	1430.71	
3/23/2009	60.90	0.00	65	0.00	47.02	0.00	0.00	1383.69	47.35
3/24/2009	34.40	34.40	65	87.19	39.31	37.57	0.00	1393.99	53.38
3/25/2009	34.00	34.00	65	87.19	31.61	37.14	0.00	1412.43	59.89
3/26/2009	34.00	34.00	65	87.19	48.91	37.14	0.00	1413.57	44.26
3/27/2009	34.00	34.00	65	87.19	39.52	37.14	0.00	1424.10	
3/28/2009	34.00	34.00	65	87.19	44.95	37.14	0.00	1429.19	
3/29/2009	34.00	34.00	65	87.19	50.39	37.14	0.00	1428.85	
3/30/2009	34.00	0.00	65	0.00	55.83	0.00	0.00	1373.02	
3/31/2009	35.78	35.78	65	271.55	61.24	39.08	0.00	1544.25	39.61
4/1/2009	39.10	39.10	65	67.61	56.31	42.72	0.00	1512.83	
4/2/2009	53.50	53.50	65	67.61	51.39	58.44	0.00	1470.62	
4/3/2009	53.00	53.00	65	67.61	46.46	57.90	0.00	1433.88	
4/4/2009	53.00	53.00	65	67.61	41.53	57.90	0.00	1402.06	
4/5/2009	53.00	53.00	65	67.61	36.60	57.90	0.00	1375.18	
4/6/2009	53.00	53.00	65	67.61	31.67	57.90	0.00	1353.23	
4/7/2009	53.00	53.00	65	67.61	26.72	57.90	0.00	1336.23	66.73
4/8/2009	53.00	53.00	65	75.20	31.65	57.90	0.00	1321.88	
4/9/2009	53.00	53.00	65	82.79	36.58	57.90	0.00	1310.20	50.69
4/10/2009	53.00	53.00	65	90.38	41.51	57.90	0.00	1301.17	
4/11/2009	53.00	53.00	65	97.96	46.44	57.90	0.00	1294.80	
4/12/2009	53.00	53.00	65	105.55	51.36	57.90	0.00	1291.08	
4/13/2009	53.00	53.00	65	113.12	56.29	57.90	0.00	1290.00	
4/14/2009	53.00	53.00	65	125.09	61.22	57.90	0.00	1295.97	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas m3/day	Estimated Biogas m3/day	Estimated Biogas Methane Content %	Influent VS Mass kg VS	Effluent VS Mass kg VS	Biogas VS Equivalent kg VS	Mass VS Wasted kg VS	Estimated MLVS kg VS	Measured SRT Days
4/15/2009	53.00	53.00	65	137.05	66.15	57.90	0.00	1308.97	
4/16/2009	53.00	53.00	65	149.02	71.08	57.90	0.00	1329.01	
4/17/2009	53.00	53.00	65	161.01	76.03	57.90	0.00	1356.09	63.80
4/18/2009	53.00	53.00	65	157.14	73.76	57.90	0.00	1381.57	
4/19/2009	53.00	53.00	65	157.14	71.49	57.90	0.00	1409.32	
4/20/2009	53.00	53.00	65	157.14	69.22	57.90	0.00	1439.34	67.64
4/21/2009	53.00	53.00	65	157.14	66.94	57.90	0.00	1471.63	
4/22/2009	53.00	53.00	65	157.14	60.13	57.90	0.00	1510.74	45.74
4/23/2009	84.60	84.60	65	160.44	60.13	92.42	0.00	1518.64	
4/24/2009	88.94	88.94	65	163.75	60.13	97.16	0.00	1525.10	63.42
4/25/2009	89.00	89.00	65	167.06	60.13	97.23	0.00	1534.80	
4/26/2009	89.00	89.00	65	170.36	60.13	97.23	0.00	1547.81	
4/27/2009	89.00	89.00	65	173.67	60.13	97.23	0.00	1564.12	
4/28/2009	93.18	93.18	65	176.98	60.13	101.80	0.00	1579.18	
4/29/2009	83.27	83.27	65	180.28	60.13	90.97	0.00	1608.36	
4/30/2009	96.78	96.78	65	183.59	60.13	105.73	0.00	1626.10	
5/1/2009	78.56	78.56	65	186.90	60.13	85.82	0.00	1667.04	
5/2/2009	79.00	79.00	65	186.90	60.13	86.30	0.00	1707.51	
5/3/2009	79.00	79.00	65	186.90	60.13	86.30	0.00	1747.97	
5/4/2009	79.89	79.89	65	186.90	60.13	87.27	0.00	1787.47	
5/5/2009	76.43	76.43	65	186.90	60.13	83.49	0.00	1830.74	
5/6/2009	76.00	76.00	65	190.27	59.93	83.03	0.00	1878.06	42.86
5/7/2009	94.29	94.29	65	182.62	59.93	103.00	0.00	1897.75	
5/8/2009	90.00	90.00	65	174.97	59.93	98.32	0.00	1914.47	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas m3/day	Estimated Biogas m3/day	Estimated Biogas Methane Content %	Influent VS Mass kg VS	Effluent VS Mass kg VS	Biogas VS Equivalent kg VS	Mass VS Wasted kg VS	Estimated MLVS kg VS	Measured SRT Days
5/9/2009	90.00	90.00	65	167.32	59.93	98.32	0.00	1923.55	
5/10/2009	90.00	90.00	65	159.67	59.93	98.32	0.00	1924.98	
5/11/2009	90.00	90.00	65	152.03	59.93	98.32	0.00	1918.76	
5/12/2009	89.67	89.67	65	144.38	59.93	97.96	0.00	1905.25	
5/13/2009	90.00	90.00	65	136.73	59.93	98.32	0.00	1883.73	
5/14/2009	96.44	96.44	65	129.08	59.93	105.36	0.00	1847.53	
5/15/2009	96.41	96.41	65	121.45	53.76	105.32	0.00	1809.91	46.89
5/16/2009	86.00	86.00	65	121.45	51.18	93.95	0.00	1786.23	
5/17/2009	86.00	86.00	65	121.45	48.61	93.95	0.00	1765.13	
5/18/2009	86.15	86.15	65	121.45	46.03	94.12	0.00	1746.44	
5/19/2009	73.36	73.36	65	121.45	43.45	80.14	0.00	1744.30	
5/20/2009	66.88	66.88	65	121.45	40.88	73.06	0.00	1751.82	
5/21/2009	68.72	68.72	65	113.56	38.36	75.07	0.00	1751.95	51.98
5/22/2009	73.05	73.05	65	113.56	40.94	79.80	0.00	1744.77	
5/23/2009	73.00	73.00	65	113.56	43.51	79.75	0.00	1735.07	
5/24/2009	73.00	73.00	65	113.56	46.09	79.75	0.00	1722.79	
5/25/2009	73.00	73.00	65	113.56	48.67	79.75	0.00	1707.94	
5/26/2009	59.72	59.72	65	113.56	51.24	65.24	0.00	1705.02	
5/27/2009	77.52	77.52	65	113.56	53.82	84.68	0.00	1680.08	
5/28/2009	89.07	89.07	65	209.27	56.36	97.30	0.00	1735.70	45.13
5/29/2009	73.00	73.00	65	140.16	54.73	79.75	0.00	1741.37	
5/30/2009	89.00	89.00	65	140.16	53.11	97.23	0.00	1731.19	
5/31/2009	89.00	89.00	65	140.16	51.49	97.23	0.00	1722.64	
6/1/2009	46.00	46.00	65	70.00	49.86	50.25	0.00	1692.52	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas m3/day	Estimated Biogas m3/day	Estimated Biogas Methane Content %	Influent VS Mass kg VS	Effluent VS Mass kg VS	Biogas VS Equivalent kg VS	Mass VS Wasted kg VS	Estimated MLVS kg VS	Measured SRT Days
6/2/2009	46.30	46.30	65	140.16	48.24	50.58	0.00	1733.86	
6/3/2009	89.00	89.00	65	140.16	46.62	97.23	0.00	1730.17	
6/4/2009	89.00	89.00	65	140.16	45.04	97.23	0.00	1728.07	52.35
6/5/2009	89.00	89.00	65	146.89	45.04	97.23	0.00	1732.70	
6/6/2009	84.00	84.00	65	146.89	45.04	91.76	0.00	1742.79	
6/7/2009	84.00	84.00	65	146.89	45.04	91.76	0.00	1752.88	
6/8/2009	83.97	83.97	65	146.89	45.04	91.73	0.00	1763.01	
6/9/2009	84.00	84.00	65	146.89	45.04	91.76	0.00	1773.11	
6/10/2009	84.00	84.00	65	146.89	45.04	91.76	0.00	1783.20	
6/11/2009	84.00	110.17	65	146.89	70.96	120.35	0.00	1738.78	41.65
6/12/2009	84.00	92.78	65	123.71	45.04	101.36	0.00	1716.09	
6/13/2009	84.00	84.00	65	123.71	45.04	91.76	0.00	1703.00	
6/14/2009	84.00	84.00	65	123.71	45.04	91.76	0.00	1689.90	
6/15/2009	84.00	84.00	65	123.71	45.04	91.76	0.00	1676.81	
6/16/2009	84.00	84.00	65	123.71	45.04	91.76	0.00	1663.71	
6/17/2009	92.51	92.51	65	123.71	45.04	101.06	0.00	1641.32	
6/18/2009	101.69	101.69	65	123.71	72.62	111.10	0.00	1581.31	39.28
6/19/2009	132.07	98.81	65	120.50	26.09	107.94	0.00	1567.78	
6/20/2009	75.00	75.00	65	117.30	26.09	81.93	0.00	1577.05	
6/21/2009	75.00	75.00	65	114.09	26.09	81.93	0.00	1583.12	
6/22/2009	75.00	75.00	65	110.88	26.09	81.93	0.00	1585.98	
6/23/2009	75.00	75.00	65	107.68	26.09	81.93	0.00	1585.64	
6/24/2009	75.00	75.00	65	104.47	26.09	81.93	0.00	1582.10	
6/25/2009	42.00	42.00	65	101.27	26.09	45.88	0.00	1611.40	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
6/26/2009	42.00	42.00	65	98.06	26.09	45.88	0.00	1637.49	
6/27/2009	42.00	42.00	65	94.86	26.09	45.88	0.00	1660.38	
6/28/2009	42.00	42.00	65	91.65	26.09	45.88	0.00	1680.06	
6/29/2009	42.00	42.00	65	88.45	26.09	45.88	84.00	1612.53	
6/30/2009	42.00	42.00	65	85.24	26.09	45.88	0.00	1625.81	
7/1/2009	42.00	42.00	65	85.20	26.09	45.88	0.00	1639.04	
7/2/2009	42.00	42.00	65	91.78	26.09	45.88	0.00	1658.84	
7/3/2009	42.00	42.00	65	98.35	26.09	45.88	0.00	1685.22	
7/4/2009	42.00	42.00	65	104.92	26.09	45.88	0.00	1718.17	
7/5/2009	42.00	42.00	65	111.49	26.09	45.88	0.00	1757.69	
7/6/2009	42.00	42.00	65	118.07	26.09	45.88	0.00	1803.79	
7/7/2009	43.68	43.68	65	124.64	26.09	47.72	0.00	1854.62	
7/8/2009	50.00	50.00	65	131.21	26.09	54.62	0.00	1905.12	
7/9/2009	70.43	70.43	65	137.83	26.21	76.94	0.00	1939.80	
7/10/2009	173.57	65.89	65	129.20	26.21	71.98	0.00	1970.81	
7/11/2009	0.00	61.50	65	120.58	26.21	67.18	0.00	1998.00	
7/12/2009	0.00	57.10	65	111.96	26.21	62.38	0.00	2021.37	
7/13/2009	0.00	52.70	65	103.34	26.21	57.57	0.00	2040.92	
7/14/2009	0.00	48.31	65	94.72	26.21	52.77	0.00	2056.66	
7/15/2009	0.00	43.91	65	86.10	26.21	47.97	0.00	2068.58	
7/16/2009	99.89	39.52	65	77.49	28.08	43.18	0.00	2074.82	
7/17/2009	103.23	42.99	65	84.29	28.08	46.96	0.00	2084.07	
7/18/2009	0.00	0.00	65	91.09	28.08	0.00	0.00	2147.08	
7/19/2009	0.00	0.00	65	97.88	28.08	0.00	0.00	2216.89	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
7/20/2009	58.22	58.22	65	103.42	27.74	63.60	233.36	1995.61	
7/21/2009	38.63	38.63	65	110.14	27.74	42.20	0.00	2035.81	
7/22/2009	82.76	82.76	65	116.85	27.74	90.41	0.00	2034.51	
7/23/2009	80.00	80.00	65	123.62	22.63	87.39	0.00	2048.11	
7/24/2009	80.00	80.00	65	124.02	22.63	87.39	0.00	2062.11	
7/25/2009	78.00	78.00	65	124.02	22.63	85.21	0.00	2078.30	
7/26/2009	78.00	78.00	65	124.02	22.63	85.21	0.00	2094.48	
7/27/2009	77.79	0.00	65	0.00	22.55	0.00	0.00	2071.94	
7/28/2009	0.00	0.00	65	0.00	22.55	0.00	0.00	2049.39	
7/29/2009	34.22	34.22	65	123.57	22.55	37.38	0.00	2113.04	
7/30/2009	50.34	50.34	65	121.50	22.43	54.99	0.00	2157.12	
7/31/2009	86.51	86.51	65	118.76	22.43	94.50	0.00	2158.95	
8/1/2009	64.00	64.00	65	116.03	22.43	69.92	0.00	2182.64	
8/2/2009	64.00	64.00	65	113.29	22.43	69.92	0.00	2203.58	
8/3/2009	63.90	63.90	65	110.10	22.33	69.81	0.00	2221.54	
8/4/2009	73.45	73.45	65	107.38	22.33	80.24	0.00	2226.34	
8/5/2009	91.58	91.58	65	104.67	93.87	100.04	0.00	2137.09	34.75
8/6/2009	50.00	50.00	65	104.67	31.90	54.62	0.00	2155.24	
8/7/2009	52.13	52.13	65	104.67	31.90	56.95	0.00	2171.07	
8/8/2009	50.00	50.00	65	104.67	31.90	54.62	0.00	2189.22	
8/9/2009	50.00	50.00	65	104.67	31.90	54.62	0.00	2207.37	
8/10/2009	50.00	50.00	65	104.31	31.79	54.62	0.00	2225.26	
8/11/2009	85.92	85.92	65	104.31	31.79	93.87	0.00	2203.91	
8/12/2009	97.42	97.42	65	104.31	31.79	106.42	0.00	2170.01	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
8/13/2009	97.00	97.00	65	104.31	31.79	105.97	0.00	2136.56	
8/14/2009	97.00	97.00	65	104.31	31.79	105.97	0.00	2103.11	
8/15/2009	97.00	97.00	65	104.31	31.79	105.97	0.00	2069.66	
8/16/2009	97.00	97.00	65	104.31	31.79	105.97	0.00	2036.22	
8/17/2009	96.63	96.63	65	111.51	33.99	105.56	0.00	2008.18	
8/18/2009	74.00	74.00	65	111.51	33.99	80.84	0.00	2004.87	
8/19/2009	37.97	73.60	65	111.51	33.99	80.40	0.00	2001.99	75.13
8/20/2009	73.83	73.83	65	111.51	33.99	80.65	0.00	1998.87	
8/21/2009	74.00	63.56	65	111.51	33.99	69.44	0.00	2006.96	
8/22/2009	74.00	74.00	65	111.51	33.99	80.84	0.00	2003.64	
8/23/2009	74.00	74.00	65	111.51	33.99	80.84	0.00	2000.33	
8/24/2009	64.22	64.22	65	111.51	33.99	70.16	0.00	2007.70	
8/25/2009	74.00	74.00	65	111.51	33.99	80.84	0.00	2004.38	
8/26/2009	74.00	49.64	65	103.42	24.28	54.23	0.00	2029.29	
8/27/2009	74.00	74.00	65	112.91	24.28	80.84	0.00	2037.08	
8/28/2009	74.00	74.00	65	122.40	24.28	80.84	0.00	2054.36	
8/29/2009	74.00	74.00	65	131.89	24.28	80.84	0.00	2081.13	
8/30/2009	74.00	74.00	65	141.39	24.28	80.84	0.00	2117.40	
8/31/2009	78.14	78.14	65	150.88	24.28	85.37	0.00	2158.63	
9/1/2009	78.00	73.77	65	160.37	24.28	80.59	0.00	2214.13	
9/2/2009	78.00	78.00	65	169.99	16.46	85.21	0.00	2282.45	
9/3/2009	80.00	80.00	65	164.88	40.00	87.39	0.00	2319.94	
9/4/2009	104.43	104.43	65	159.78	40.00	114.08	0.00	2325.64	
9/5/2009	85.00	85.00	65	154.67	40.00	92.86	0.00	2347.45	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
9/6/2009		85.00	65	149.57	40.00	92.86	0.00	2364.16	
9/7/2009		85.00	65	144.46	40.00	92.86	0.00	2375.76	
9/8/2009		84.66	65	139.35	40.00	92.48	0.00	2382.63	
9/9/2009		85.00	65	134.25	40.00	92.86	0.00	2384.02	
9/10/2009		85.00	65	129.14	40.00	92.86	0.00	2380.30	
9/11/2009		83.03	65	123.86	81.66	90.71	0.00	2331.80	40.45
9/12/2009		80.00	65	133.59	67.00	87.39	0.00	2311.00	
9/13/2009		80.00	65	143.32	67.00	87.39	0.00	2299.92	
9/14/2009		92.03	65	164.34	60.00	100.54	0.00	2303.72	
9/15/2009		45.92	65	82.00	60.00	50.16	0.00	2275.56	
9/16/2009		45.92	65	82.00	60.00	50.16	0.00	2247.39	53.97
9/17/2009		104.31	65	186.27	60.00	113.95	0.00	2259.71	
9/18/2009		99.09	65	176.94	60.00	108.25	0.00	2268.41	
9/19/2009		93.86	65	167.62	60.00	102.54	0.00	2273.48	
9/20/2009		88.64	65	158.29	60.00	96.83	0.00	2274.93	
9/21/2009		83.42	65	148.96	60.00	91.13	0.00	2272.76	
9/22/2009		47.92	65	85.58	28.00	52.35	0.00	2277.99	
9/23/2009		84.55	65	150.97	40.00	92.36	0.00	2296.60	
9/24/2009		84.55	65	150.97	40.00	92.36	0.00	2315.21	
9/25/2009		84.55	65	150.97	40.00	92.36	0.00	2333.83	
9/26/2009		84.55	65	150.97	40.00	92.36	0.00	2352.44	
9/27/2009		84.55	65	150.97	40.00	92.36	0.00	2371.05	
9/28/2009		91.90	65	164.10	40.00	100.39	0.00	2394.76	
9/29/2009		91.90	65	164.10	40.00	100.39	0.00	2418.47	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas m3/day	Estimated Biogas m3/day	Estimated Biogas Methane Content %	Influent VS Mass kg VS	Effluent VS Mass kg VS	Biogas VS Equivalent kg VS	Mass VS Wasted kg VS	Estimated MLVS kg VS	Measured SRT Days
9/30/2009		0.00	65	0.00	40.30	0.00	0.00	2378.18	
10/1/2009		91.90	65	164.10	57.00	100.39	0.00	2384.89	
10/2/2009		91.90	65	164.10	57.00	100.39	0.00	2391.60	
10/3/2009		91.90	65	164.10	57.00	100.39	0.00	2398.30	
10/4/2009		91.90	65	164.10	57.00	100.39	0.00	2405.01	
10/5/2009		100.58	65	179.62	44.11	109.88	0.00	2430.64	
10/6/2009		100.58	65	179.62	44.11	109.88	0.00	2456.27	
10/7/2009		154.60	65	276.08	139.16	168.89	0.00	2424.29	28.10
10/8/2009		91.84	65	164.00	92.22	100.33	0.00	2395.74	
10/9/2009		91.84	65	164.00	92.22	100.33	0.00	2367.19	
10/10/2009		91.84	65	164.00	92.22	100.33	0.00	2338.64	
10/11/2009		91.84	65	164.00	92.22	100.33	0.00	2310.09	
10/12/2009		91.84	65	164.00	101.89	100.33	0.00	2271.86	
10/13/2009		91.84	65	164.00	101.89	100.33	0.00	2233.64	
10/14/2009		131.22	65	234.32	101.89	143.35	0.00	2222.72	37.45
10/15/2009		99.12	65	177.00	66.64	108.28	0.00	2224.81	
10/16/2009		88.50	65	177.00	66.64	96.68	0.00	2238.49	
10/17/2009		88.50	65	177.00	66.64	96.68	0.00	2252.18	
10/18/2009		88.50	65	177.00	66.64	96.68	0.00	2265.86	
10/19/2009		88.50	65	177.00	66.64	96.68	0.00	2279.54	
10/20/2009		88.50	65	177.00	66.64	96.68	0.00	2293.23	
10/21/2009		88.50	65	177.00	66.64	96.68	0.00	2306.91	
10/22/2009		88.50	65	177.00	66.64	96.68	0.00	2320.60	
10/23/2009		140.04	65	280.08	66.64	152.99	0.00	2381.06	55.26

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
10/24/2009		88.38	65	176.77	66.64	96.55	0.00	2394.64	0.00
10/25/2009		66.50	65	133.00	66.64	72.65	0.00	2388.35	
10/26/2009		66.50	65	133.00	60.00	72.65	0.00	2388.71	
10/27/2009		66.50	65	133.00	60.00	72.65	0.00	2389.06	
10/28/2009		66.50	65	133.00	60.00	72.65	0.00	2389.41	
10/29/2009		66.50	65	133.00	60.00	72.65	0.00	2389.77	
10/30/2009		66.50	65	133.00	60.00	72.65	0.00	2390.12	
10/31/2009		66.50	65	133.00	60.00	72.65	0.00	2390.47	
11/1/2009		66.50	65	133.00	60.00	72.65	0.00	2390.82	
11/2/2009		66.50	65	133.00	60.00	72.65	0.00	2391.18	
11/3/2009		66.50	65	133.00	60.00	72.65	0.00	2391.53	
11/4/2009		66.50	65	133.00	60.00	72.65	0.00	2391.88	
11/5/2009		66.50	65	133.00	60.00	72.65	0.00	2392.24	
11/6/2009		66.50	65	133.00	89.22	72.65	0.00	2363.37	
11/7/2009		62.51	65	133.00	66.64	68.29	0.00	2361.44	
11/8/2009		62.51	65	133.00	66.64	68.29	0.00	2359.52	
11/9/2009		52.26	65	111.18	41.91	57.09	0.00	2371.70	
11/10/2009		62.42	65	132.81	50.07	68.19	0.00	2386.26	
11/11/2009		72.59	65	154.44	58.22	79.30	0.00	2403.18	
11/12/2009		82.76	65	176.07	66.37	90.40	0.00	2422.48	
11/13/2009		42.64	65	90.72	74.53	46.58	0.00	2392.09	
11/14/2009		47.94	65	102.00	74.53	52.37	0.00	2367.19	
11/15/2009		47.94	65	102.00	74.53	52.37	0.00	2342.29	
11/16/2009		47.94	65	102.00	74.53	52.37	0.00	2317.39	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
11/17/2009		47.97	65	102.06	83.84	52.40	0.00	2283.21	
11/18/2009		47.97	65	102.06	83.84	52.40	0.00	2249.02	
11/19/2009		47.97	65	102.06	83.84	52.40	0.00	2214.83	
11/20/2009		47.97	65	102.06	83.84	52.40	0.00	2180.65	
11/21/2009		47.97	65	102.06	83.84	52.40	0.00	2146.46	
11/22/2009		47.97	65	102.06	83.84	52.40	0.00	2112.28	
11/23/2009		47.97	65	102.06	83.84	52.40	0.00	2078.09	
11/24/2009		54.82	65	116.64	95.82	59.89	0.00	2039.02	
11/25/2009		54.82	65	116.64	95.82	59.89	0.00	1999.95	
11/26/2009		54.82	65	116.64	95.82	59.89	0.00	1960.88	
11/27/2009		54.82	65	116.64	95.82	59.89	0.00	1921.81	
11/28/2009		54.82	65	116.64	95.82	59.89	0.00	1882.74	
11/29/2009		54.82	65	116.64	95.82	59.89	0.00	1843.67	
11/30/2009		54.82	65	116.64	95.82	59.89	0.00	1804.60	
12/1/2009		54.82	65	116.64	95.82	59.89	0.00	1765.53	
12/2/2009		65.85	65	140.10	57.55	71.93	0.00	1776.15	83.16
12/3/2009		64.54	65	137.31	57.55	70.50	0.00	1785.41	
12/4/2009		63.22	65	134.52	57.55	69.07	0.00	1793.32	
12/5/2009		61.91	65	131.73	57.55	67.64	0.00	1799.86	
12/6/2009		60.60	65	128.94	57.55	66.20	0.00	1805.05	
12/7/2009		72.51	65	154.27	70.38	79.21	0.00	1809.73	
12/8/2009		70.90	65	150.86	70.38	77.46	0.00	1812.76	
12/9/2009		69.30	65	147.45	70.38	75.71	0.00	1814.12	
12/10/2009		67.70	65	144.03	70.38	73.95	0.00	1813.82	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
12/11/2009		66.09	65	140.62	70.38	72.20	0.00	1811.86	
12/12/2009		64.49	65	137.21	70.38	70.45	0.00	1808.24	
12/13/2009		62.88	65	133.79	70.38	68.70	0.00	1802.96	
12/14/2009		62.98	65	134.00	70.38	68.80	0.00	1797.78	
12/15/2009		62.98	65	134.00	70.38	68.80	0.00	1792.60	
12/16/2009		62.98	65	134.00	59.46	68.80	0.00	1798.34	56.86
12/17/2009		62.98	65	134.00	70.38	68.80	0.00	1793.16	
12/18/2009		58.39	65	124.24	70.38	63.79	0.00	1783.23	
12/19/2009		58.39	65	124.24	70.38	63.79	0.00	1773.30	
12/20/2009		58.39	65	124.24	70.38	63.79	0.00	1763.37	
12/21/2009		58.39	65	124.24	70.38	63.79	0.00	1753.44	
12/22/2009		58.39	65	124.24	70.38	63.79	0.00	1743.51	
12/23/2009		58.39	65	124.24	70.38	63.79	0.00	1733.58	
12/24/2009		58.39	65	124.24	70.38	63.79	0.00	1723.65	
12/25/2009		58.39	65	124.24	70.38	63.79	0.00	1713.72	
12/26/2009		58.39	65	124.24	70.38	63.79	0.00	1703.79	
12/27/2009		58.39	65	124.24	70.38	63.79	0.00	1693.86	
12/28/2009		58.39	65	124.24	70.38	63.79	0.00	1683.93	
12/29/2009		58.39	65	124.24	70.38	63.79	0.00	1674.00	
12/30/2009		58.39	65	124.24	70.38	63.79	0.00	1664.07	
12/31/2009		58.39	65	124.24	70.38	63.79	0.00	1654.14	
1/1/2010		58.39	65	124.24	70.38	63.79	0.00	1644.21	
1/2/2010		58.39	65	124.24	70.38	63.79	0.00	1634.28	
1/3/2010		58.39	65	124.24	70.38	63.79	0.00	1624.35	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
1/4/2010		58.39	65	124.24	70.38	63.79	0.00	1614.42	
1/5/2010		58.39	65	124.24	70.38	63.79	0.00	1604.49	
1/6/2010		58.39	65	124.24	70.38	63.79	0.00	1594.56	
1/7/2010		58.39	65	124.24	70.38	63.79	0.00	1584.63	
1/8/2010		58.39	65	124.24	70.38	63.79	0.00	1574.70	
1/9/2010		58.39	65	124.24	70.38	63.79	0.00	1564.77	
1/10/2010		58.39	65	124.24	70.38	63.79	0.00	1554.84	
1/11/2010		58.39	65	124.24	70.38	63.79	155.48	1389.43	
1/12/2010		58.39	65	124.24	70.38	63.79	0.00	1379.50	
1/13/2010		58.39	65	124.24	70.38	63.79	0.00	1369.57	
1/14/2010		58.39	65	124.24	70.38	63.79	0.00	1359.64	
1/15/2010		58.39	65	124.24	70.38	63.79	0.00	1349.71	
1/16/2010		58.39	65	124.24	70.38	63.79	0.00	1339.78	
1/17/2010		58.39	65	124.24	70.38	63.79	0.00	1329.85	
1/18/2010		58.39	65	124.24	70.38	63.79	0.00	1319.92	
1/19/2010		58.39	65	124.24	70.38	63.79	0.00	1309.99	
1/20/2010		58.39	65	124.24	70.38	63.79	0.00	1300.06	
1/21/2010		58.39	65	124.24	70.38	63.79	0.00	1290.13	
1/22/2010		58.39	65	124.24	70.38	63.79	0.00	1280.20	
1/23/2010		58.39	65	124.24	70.38	63.79	0.00	1270.27	
1/24/2010		58.39	65	124.24	70.38	63.79	0.00	1260.34	
1/25/2010		58.28	65	124.00	87.00	63.67	0.00	1233.67	28.55
1/26/2010		70.74	65	150.52	86.97	77.28	0.00	1219.94	
1/27/2010		71.16	65	151.40	86.97	77.74	0.00	1206.64	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

Date	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
1/28/2010		71.58	65	152.29	86.97	78.19	0.00	1193.77	
1/29/2010		71.99	65	153.18	86.97	78.65	0.00	1181.34	
1/30/2010		72.41	65	154.07	86.97	79.10	0.00	1169.34	
1/31/2010		72.83	65	154.95	86.97	79.56	0.00	1157.76	
2/1/2010		73.25	65	155.84	86.97	80.02	0.00	1146.62	
2/2/2010		73.66	65	156.73	86.97	80.47	0.00	1135.92	
2/3/2010		74.08	65	157.62	86.97	80.93	0.00	1125.64	
2/4/2010		74.50	65	158.50	86.97	81.38	0.00	1115.79	
2/5/2010		74.91	65	159.39	86.97	81.84	0.00	1106.38	
2/6/2010		75.33	65	160.28	86.97	82.29	0.00	1097.40	
2/7/2010		75.75	65	161.17	86.97	82.75	0.00	1088.85	
2/8/2010		76.16	65	162.05	86.97	83.21	0.00	1080.73	
2/9/2010		76.58	65	162.94	86.97	83.66	0.00	1073.05	
2/10/2010		77.00	65	163.83	86.97	84.12	0.00	1065.79	
2/11/2010		77.42	65	164.72	86.97	84.57	0.00	1058.97	
2/12/2010		77.83	65	165.60	86.97	85.03	0.00	1052.58	
2/13/2010		78.25	65	166.49	86.97	85.48	0.00	1046.62	
2/14/2010		78.67	65	167.38	86.97	85.94	0.00	1041.09	
2/15/2010		79.08	65	168.26	86.97	86.39	0.00	1036.00	
2/16/2010		79.50	65	169.15	86.97	86.85	0.00	1031.34	
2/17/2010		79.92	65	170.04	86.97	87.31	0.00	1027.10	
2/18/2010		80.34	65	170.93	86.97	87.76	0.00	1023.30	
2/19/2010		80.75	65	171.81	86.97	88.22	0.00	1019.94	
2/20/2010		81.17	65	172.70	86.97	88.67	0.00	1017.00	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
2/21/2010		81.59	65	173.59	86.97	89.13	0.00	1014.50	
2/22/2010		82.00	65	174.48	86.97	89.58	0.00	1012.42	
2/23/2010		82.42	65	175.36	86.97	90.04	0.00	1010.78	
2/24/2010		82.84	65	176.25	86.97	90.50	0.00	1009.57	
2/25/2010		83.26	65	177.14	86.97	90.95	0.00	1008.79	
2/26/2010		83.67	65	178.03	86.97	91.41	0.00	1008.45	
2/27/2010		84.09	65	178.91	86.97	91.86	0.00	1008.53	
2/28/2010		84.51	65	179.80	86.97	92.32	0.00	1009.05	
3/1/2010		84.92	65	180.69	86.97	92.77	0.00	1010.00	
3/2/2010		85.34	65	181.58	86.97	93.23	0.00	1011.38	
3/3/2010		85.76	65	182.46	86.97	93.68	0.00	1013.20	
3/4/2010		86.17	65	183.35	86.97	94.14	0.00	1015.44	
3/5/2010		86.59	65	184.24	86.97	94.60	0.00	1018.12	
3/6/2010		87.01	65	185.13	86.97	95.05	0.00	1021.23	
3/7/2010		87.43	65	186.01	86.97	95.51	0.00	1024.77	
3/8/2010		87.84	65	186.90	86.97	95.96	0.00	1028.74	
3/9/2010		88.26	65	187.79	86.97	96.42	0.00	1033.14	
3/10/2010		88.68	65	188.68	86.97	96.87	0.00	1037.98	
3/11/2010		89.09	65	189.56	86.97	97.33	0.00	1043.24	
3/12/2010		89.51	65	190.45	86.97	97.79	0.00	1048.94	
3/13/2010		89.93	65	191.34	86.97	98.24	0.00	1055.07	
3/14/2010		90.35	65	192.22	86.97	98.70	0.00	1061.64	
3/15/2010		90.76	65	193.11	86.97	99.15	0.00	1068.63	
3/16/2010		91.18	65	194.00	86.97	99.61	0.00	1076.06	

Appendix 5. OSU SREC ASBR organic matter mass balance data continued

	Measured Biogas	Estimated Biogas	Estimated Biogas Methane Content	Influent VS Mass	Effluent VS Mass	Biogas VS Equivalent	Mass VS Wasted	Estimated MLVS	Measured SRT
Date	m3/day	m3/day	%	kg VS	kg VS	kg VS	kg VS	kg VS	Days
3/17/2010		91.60	65	194.89	86.97	100.06	0.00	1083.91	
3/18/2010		92.01	65	195.77	86.97	100.52	0.00	1092.20	
3/19/2010		92.43	65	196.66	86.97	100.98	0.00	1100.93	
3/20/2010		92.85	65	197.55	86.97	101.43	0.00	1110.08	
3/21/2010		93.27	65	198.44	86.97	101.89	0.00	1119.66	
3/22/2010		93.68	65	199.32	86.97	102.34	0.00	1129.68	
3/23/2010		94.10	65	200.21	86.97	102.80	0.00	1140.13	
3/24/2010		94.52	65	201.10	86.97	103.25	0.00	1151.01	
3/25/2010		94.93	65	201.99	86.97	103.71	0.00	1162.32	
3/26/2010		95.35	65	202.87	86.97	104.16	0.00	1174.06	
3/27/2010		95.77	65	203.76	86.97	104.62	0.00	1186.24	
3/28/2010		96.18	65	204.65	86.97	105.08	0.00	1198.85	
3/29/2010		96.60	65	205.54	86.97	105.53	0.00	1211.89	
3/30/2010		97.02	65	206.42	86.97	105.99	0.00	1225.36	
3/31/2010		97.44	65	207.31	86.97	106.44	0.00	1239.26	
4/1/2010		97.85	65	208.20	86.97	106.90	0.00	1253.59	
4/2/2010		98.27	65	209.09	86.97	107.35	0.00	1268.36	
4/3/2010		98.69	65	209.97	86.97	107.81	0.00	1283.56	
4/4/2010		99.10	65	210.86	86.97	108.27	0.00	1299.19	
4/5/2010		99.52	65	211.75	86.97	108.72	0.00	1315.25	
4/6/2010		99.94	65	212.63	86.97	109.18	0.00	1331.74	
4/13/2010		100.36	65	213.52	86.97	109.63	0.00	1348.67	
4/22/2010		101.35	65	215.64	86.97	110.72	0.00	1366.62	

Appendix 6. Swine Manure Total and Volatile Solids for Table 11

Date	Total Solids mg / l	Volatile Solids mg / l
10/15/2010	8,093	5,723
10/21/2010	8,093	5,723
10/29/2010	5,217	3,760
11/3/2010	3,339	2,011
11/11/2010	7,612	4,691
11/17/2010	3,450	2,078
11/29/2010	4,426	2,706
12/1/2010	3,640	2,059
12/6/2010	6,987	4,638
12/10/2010	7,546	5,042
12/13/2010	7,899	5,452
12/15/2010	27,873	22,963
1/13/2011	14,687	11,507
1/18/2011	5,656	3,982
1/20/2011	5,843	4,194
1/25/2011	4,993	3,431
1/27/2011	7,722	5,862
2/8/2011	4,489	3,071
2/15/2011	14,660	11,348
2/17/2011	4,904	3,366
2/22/2011	3,842	2,457
2/24/2011	3,679	2,383
3/1/2011	5,902	4,259
3/3/2011	5,864	4,208
3/8/2011	3,691	2,471
3/10/2011	3,867	2,596
3/15/2011	5,244	3,653
3/17/2011	5,034	3,563
3/22/2011	8,778	6,714
3/29/2011	4,524	3,016
3/31/2011	4,962	3,437
4/7/2011	3,579	2,498

Appendix 6. Swine Manure Total and Volatile Solids for Table 11continued

Date	Total Solids mg / l	Volatile Solids mg / l
4/12/2011	3,353	2,332
4/14/2011	3,696	2,469
4/19/2011	5,219	3,552
4/21/2011	4,421	3,028
4/26/2011	4,380	2,791
5/3/2011	9,323	7,156
5/10/2011	8,799	6,360
5/17/2011	8,087	5,733
5/19/2011	19,753	14,186
5/27/2011	8,193	5,900
5/31/2011	9,927	7,438
6/8/2011	10,326	7,414
6/13/2011	7,429	5,177
6/16/2011	9,403	6,376
6/20/2011	14,391	10,686
6/23/2011	20,776	16,768
6/27/2011	12,067	9,076
11/4/2011	12,246	9,189
10/27/2011	19,898	15,546
12/1/2011	4,154	2,926
12/8/2011	7,032	5,382
1/19/2012	8,420	6,038
1/31/2012	5,099	3,497
2/7/2012	12,689	6,534
2/28/2012	11,657	8,442
3/6/2012	8,017	5,939
3/13/2012	8,277	6,167
3/27/2012	13,504	10,199
4/5/2012	17,084	13,034
4/10/2012	10,271	7,360
4/17/2012	8,439	6,164
4/24/2012	8,097	5,961
Average	8,289	5,995
Standard Deviation	4,921	3,992
n	64	64

Appendix 7. Swine Manure Characteristics for table 11

Date Sampled	Parameter					
	Chemical Oxygen Demand	pH	Total N	Phosphorus (P_2O_5)	Potassium (K_2O)	Calcium
	g/l		mg/l	mg/l	mg/l	mg/l
12/6/2010	10.37	7.1	9.13	416.8	633.2	228
1/13/2011	19.6	7	10.3	531.28	572.38	236
2/8/2011	7.61	7.3	4.3	332.05	324.15	195
3/1/2011	8.74	7.1	5.89	373.3	378.4	290
3/29/2011	7.13	7.1	5.48	332	365.1	141
4/19/2011	6.08	7.3	4.7	286.25	331.38	199
5/25/2011	8.34	7.3	6.6	572.5	250.64	332
6/8/2011	11.54	7.1	6.4	613.72	506.1	313
10/27/2011	28.73	7.6		1131.26	426.57	867
12/1/2011	4.95	8		293.12	279.56	179
12/8/2011	10.63	8.4		265.64	281.97	307
1/19/2012	10.52	8.3		425.94	409.7	353
1/31/2001	5.89	8.1	7.7	233.58	360.29	234
2/7/2012	20.5	7	11.7	739.67	502.49	305
2/21/2012	18.18	7.2	10.8	462.58	625.4	294
2/28/2012	15.81	7.5	9.4	338.92	543.46	199
3/6/2012	11.43	7.6	10.3	302.28	545.86	193
3/13/2012	11.46	8.2	8.4	135.11	456.08	141
3/27/2012	15.79	7.1	5.8	748.83	550.68	526
4/3/2012	19.41	7.8	7.2	1149.58	486.82	781
4/10/2012	11.6	8	7.7	1433.54	501.28	986
4/17/2012	8.6	7.8	8.2	689.29	360.29	523
4/24/2012	9.11	7.8	6.4	735.09	342.22	545
Average	12.26	7.55	7.71	545.32	436.26	363.78
Standard Deviation	5.83	0.45	2.13	328.44	113.24	235.16
n	23	23	19	23	23	23

Appendix 7. Swine Manure Characteristics for table 11 continued

Date Sampled	Parameter						
	Magnesium	Sodium	Sulfur	Iron	Zinc	Copper	Manganese
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
12/6/2010	106.5	176	89	10.66	6.66	1.03	2.34
1/13/2011	132	159	54	99	8.2	1.2	2.9
2/8/2011	88	99	45	9.3	7.6	1.1	2.3
3/1/2011	87	113	78	17.69	11.83	1.61	2.94
3/29/2011	87	91	38	4.63	3.8	0.59	1.7
4/19/2011	71	97	53	15	9.7	2	2.3
5/25/2011	119	115	63	19.9	14.4	2.1	5.4
6/8/2011	138	210	83	25.7	21.3	2.5	5.5
10/27/2011	208	130	154	75.5	53.5	6.1	12.9
12/1/2011	95	102	36	8	7.8	0.7	3
12/8/2011	58	102	75	23.4	22.2	2.5	4.5
1/19/2012	103	117	78	26	25	2.6	5.2
1/31/2001	56	110	55	14.3	15	1.7	3.1
2/7/2012	197	160	86	23.7	19.3	2.5	6.7
2/21/2012	118	157	86	24.2	20.9	2.6	5.1
2/28/2012	77	132	79	15.5	13.5	1.8	3.3
3/6/2012	64	132	71	12.8	10.8	1.3	2.7
3/13/2012	26	122	38	5.5	5.3	0.7	1
3/27/2012	174	142	128	25	25.8	3.4	8.7
4/3/2012	224	136	172	65.1	40.1	5.3	13.2
4/10/2012	267	132	212	87.9	52.9	7.7	16.8
4/17/2012	124	114	119	43.5	25.3	3.6	8.2
4/24/2012	123	104	130	48.3	27	4	8.6
Average	119.24	128.35	87.91	30.46	19.47	2.55	5.58
Standard Deviation	59.53	28.87	45.49	26.85	13.82	1.81	4.13
n	23	23	23	23	23	23	23

Appendix 8. Glycerol Chemical Oxygen Demand Analysis

Date Measured	Location	ID	Dilution	Absorbance	COD Unadjusted g/l	COD Adjusted g/l
8/17/2010	Tulsa	1.1	722	0.673	1.90	**1372.05**
8/17/2010	Tulsa	1.2	722	0.724	2.04	**1476.03**
8/17/2010	Tulsa	1.3	722	0.75	2.12	**1529.03**
8/17/2010	Tulsa	2.1	722	0.709	2.00	**1445.45**
8/17/2010	Tulsa	2.2	722	0.635	1.79	**1294.58**
8/17/2010	Tulsa	2.3	722	0.685	1.93	**1396.52**
8/17/2010	Tulsa	3.1	722	0.657	1.86	**1339.43**
8/17/2010	Tulsa	3.2	722	0.657	1.86	**1339.43**
8/17/2010	Tulsa	3.3	722	0.613	1.73	**1249.73**
					Average	**1,382.47**
					Standard Deviation	**89.13**

Appendix 9. Biogas analysis results of glycerol treatment and control reactors

	Glycerol Treatment Reactor				Control Reactor			
Gas Parameter	CH_4 %	CO_2 %	H_2 %	N_2 %	CH_4 %	CO_2 %	H_2 %	N_2 %
1/21/2011	74.55	21.23	0	3.58	67.78	29.2	0	2.6
3/1/2011	72.56	23.75	0	3.69	65.84	30.95	0	3.21
5/6/2011	71.29	25.45	0	3.26	64.89	32.22	0	2.89
6/13/2011	74.96	21.59	0	3.45	66.18	30.37	0	3.45
						100		
Average	73.34	23.01	0.00	3.50	66.17	30.69	0.00	3.04
Standard Deviation	1.72	1.97	0.00	0.18	1.20	1.26	0.00	0.37

Appendix 10. Control Reactor Influent Parameters

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
10/15/2010	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/21/2010	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/29/2010	8,429	5,217	3,760	1,457	3,515	2,957	803	558	899	1,702
11/3/2010	5,167	3,339	2,011	1,328	1,333	1,034	977	299	1,029	2,006
11/11/2010	12,283	7,612	4,691	2,921	3,738	2,803	1,888	935	1,986	3,874
11/17/2010	4,960	3,450	2,078	1,372	1,231	1,019	1,059	212	1,160	2,219
11/29/2010	6,203	4,426	2,706	1,720	2,035	1,582	1,124	453	1,267	2,391
12/1/2010	5,930	3,640	2,059	1,581	1,170	900	1,159	270	1,311	2,470
12/6/2010	10,372	6,987	4,638	2,349	3,899	3,142	1,496	757	1,592	3,088
12/10/2010	12,537	7,546	5,042	2,504	4,327	3,418	1,624	909	1,595	3,219
12/13/2010	12,782	7,899	5,452	2,447	4,815	3,860	1,592	955	1,492	3,084
12/15/2010	21,968	27,873	22,963	4,910	25,622	21,944	1,019	3,678	1,232	2,251
1/13/2011	19,597	14,687	11,507	3,180	12,418	10,405	1,102	2,013	1,167	2,269
1/18/2011	10,175	5,656	3,982	1,674	3,695	2,982	1,000	713	961	1,961
1/20/2011	9,987	5,843	4,194	1,649	3,624	3,036	1,158	588	1,061	2,219
1/25/2011	9,215	4,993	3,431	1,562	2,700	2,151	1,280	549	1,013	2,293
1/27/2011	12,358	7,722	5,862	1,860	5,740	4,805	1,057	935	925	1,982
2/8/2011	7,615	4,489	3,071	1,418	2,432	2,043	1,028	389	1,029	2,057
2/15/2011	15,446	14,660	11,348	3,312	12,211	10,060	1,288	2,151	1,161	2,449
2/17/2011	7,088	4,904	3,366	1,538	2,847	2,319	1,047	528	1,010	2,057
2/22/2011	6,711	3,842	2,457	1,385	1,831	1,460	997	371	1,014	2,011
2/24/2011	5,779	3,679	2,383	1,296	1,358	1,173	1,210	185	1,111	2,321
3/1/2011	8,744	5,902	4,259	1,643	3,786	3,212	1,047	574	1,069	2,116

Appendix 10. Control Reactor Influent Parameters continued

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
3/3/2011	8,471	5,864	4,208	1,656	3,616	3,018	1,190	598	1,058	2,248
3/8/2011	5,167	3,691	2,471	1,220	1,605	1,354	1,117	251	969	2,086
3/10/2011	5,535	3,867	2,596	1,271	1,923	1,636	960	287	984	1,944
3/15/2011	7,266	5,244	3,653	1,591	3,286	2,778	876	508	1,082	1,958
3/17/2011	7,342	5,034	3,563	1,471	3,051	2,572	991	479	992	1,983
3/22/2011	13,111	8,778	6,714	2,064	6,885	5,838	876	1,047	1,017	1,893
3/29/2011	7,125	4,524	3,016	1,508	2,398	1,929	1,087	469	1,039	2,126
3/31/2011	7,106	4,962	3,437	1,525	2,763	2,258	1,179	505	1,020	2,199
4/7/2011	4,527	3,579	2,498	1,081	2,003	1,722	776	281	800	1,576
4/12/2011	4,706	3,353	2,332	1,021	1,815	1,485	847	330	691	1,538
4/14/2011	4,734	3,696	2,469	1,227	1,786	1,451	1,018	335	892	1,910
4/19/2011	6,080	5,219	3,552	1,667	3,389	2,755	797	634	1,033	1,830
4/21/2011	5,346	4,421	3,028	1,393	2,649	2,205	823	444	949	1,772
4/26/2011	6,532	4,380	2,791	1,589	2,071	1,720	1,071	351	1,238	2,309
5/3/2011	13,535	9,323	7,156	2,167	7,464	6,265	891	1,199	968	1,859
5/10/2011	12,264	8,799	6,360	2,439	6,453	5,340	1,020	1,113	1,326	2,346
5/17/2011	13,140	8,087	5,733	2,354	5,885	4,776	957	1,109	1,245	2,202
5/19/2011	22,119	19,753	14,186	5,568	17,933	13,152	1,033	4,781	787	1,820
5/25/2011	8,339									
5/27/2011		8,193	5,900	2,293	5,684	4,812	1,088	872	1,421	2,509
5/31/2011		9,927	7,438	2,489	8,053	6,437	1,001	1,617	872	1,873
6/8/2011	11,540	10,326	7,414	2,911	7,565	6,064	1,350	1,501	1,410	2,760
6/13/2011	12,984	7,429	5,177	2,252	5,044	3,849	1,328	1,196	1,056	2,384

Appendix 10. Control Reactor Influent Parameters continued

	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
Date	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
6/16/2011	12,984	9,403	6,376	3,028	6,311	5,001	1,374	1,310	1,718	3,092
6/20/2011	11,822	14,391	10,686	3,706	11,263	9,148	1,538	2,116	1,590	3,128
6/23/2011	24,096	20,776	16,768	4,008	18,724	15,748	1,020	2,977	1,031	2,051
6/27/2011	24,905	12,067	9,076	2,991	9,789	8,047	1,029	1,742	1,249	2,278
6/29/2011	20,067									

Appendix 11. Control Reactor Effluent Parameters

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
10/15/2010	1,026	3,124	1,471	1,653	1,246	924	547	322	1,331	1,878
10/21/2010	2,899	3,564	1,746	1,818	1,164	760	986	404	1,414	2,400
10/29/2010	1,482	2,597	1,235	1,362	600	398	837	202	1,160	1,997
11/3/2010	1,666	2,766	1,456	1,310	962	694	762	268	1,042	1,804
11/11/2010	6,316	4,902	2,704	2,198	1,573	1,187	1,518	386	1,811	3,329
11/17/2010	2,748	3,596	1,941	1,655	914	725	1,216	189	1,466	2,682
11/29/2010	8,650	5,328	3,320	2,008	2,449	1,900	1,420	549	1,459	2,879
12/1/2010	1,318	2,388	967	1,421	509	299	668	210	1,211	1,879
12/6/2010	1,732	3,037	1,381	1,656	729	507	874	222	1,434	2,308
12/10/2010	3,228	3,670	1,778	1,892	1,019	728	1,050	291	1,601	2,651
12/13/2010	3,530	4,212	2,221	1,991	1,739	1,183	1,038	556	1,435	2,473
12/15/2010	9,808	8,897	5,627	3,270	6,843	4,894	733	1,949	1,321	2,054
1/13/2011	19,154	15,366	11,318	4,048	13,778	10,632	686	3,146	902	1,588
1/18/2011	2,908	3,390	1,797	1,593	1,477	1,015	782	462	1,131	1,913
1/20/2011	1,911	2,836	1,476	1,360	806	677	799	129	1,231	2,030
1/25/2011	2,457	2,529	1,203	1,326	625	429	774	196	1,130	1,904
1/27/2011	1,572	2,490	1,140	1,350	669	444	696	225	1,125	1,821
2/8/2011	1,195	2,368	1,133	1,235	657	492	641	165	1,070	1,711
2/15/2011	1,685	2,247	1,202	1,045	588	454	748	134	911	1,659
2/17/2011	2,268	2,874	1,473	1,401	981	760	713	221	1,180	1,893
2/22/2011	2,184	2,289	1,012	1,277	597	422	590	175	1,102	1,692
2/24/2011	2,363	2,390	1,176	1,214	631	524	652	107	1,107	1,759
3/1/2011	1,600	2,210	978	1,232	579	430	548	149	1,083	1,631

Appendix 11. Control Reactor Effluent Parameters continued

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
3/3/2011	1,280	2,323	1,072	1,251	591	440	632	151	1,100	1,732
3/8/2011	1,600	2,357	1,200	1,157	721	532	668	189	968	1,636
3/10/2011	1,242	2,069	900	1,169	589	380	520	209	960	1,480
3/15/2011	1,167	2,077	859	1,218	558	369	490	189	1,029	1,519
3/17/2011	1,261	2,131	1,007	1,124	638	458	549	180	944	1,493
3/22/2011	1,713	2,562	1,217	1,345	1,070	713	504	357	988	1,492
3/29/2011	9,186	3,084	1,680	1,404	1,693	1,137	543	556	848	1,391
3/31/2011	1,873	2,674	1,312	1,362	1,233	755	557	478	884	1,441
4/7/2011	2,099	2,950	1,613	1,337	1,036	815	798	221	1,116	1,914
4/12/2011	1,995	2,546	1,332	1,214	757	530	802	227	987	1,789
4/14/2011	1,346	2,036	1,012	1,024	684	463	549	221	803	1,352
4/19/2011	1,073	2,159	1,282	877	655	800	482	-145	1,022	1,504
4/21/2011	1,092	2,203	1,023	1,180	791	513	510	278	902	1,412
4/26/2011	1,591	2,590	1,314	1,276	922	632	682	290	986	1,668
5/3/2011	1,393	2,685	1,330	1,355	763	628	702	135	1,220	1,922
5/10/2011	1,666	2,729	1,304	1,425	955	660	644	295	1,130	1,774
5/17/2011	4,386	4,732	2,814	1,918	2,776	2,145	669	631	1,287	1,956
5/19/2011	2,664	3,169	1,774	1,394	1,206	574	1,200	631	763	1,963
5/25/2011	1,506	2,374	1,087	1,288	1,074	822	265	253	1,035	1,300
5/27/2011		2,576	1,362	1,213	1,064	776	587	289	924	1,511
5/31/2011		4,628	3,082	1,545	2,833	2,158	924	675	870	1,794
6/8/2011	6,429	8,472	6,121	2,351	7,677	5,765	357	1,912	439	796
6/13/2011	2,108	3,168	1,838	1,330	1,414	1,174	663	240	1,090	1,753

Appendix 11. Control Reactor Effluent Parameters continued

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
6/16/2011	5,422	8,645	5,983	2,662	7,196	5,426	557	1,770	892	1,449
6/20/2011	4,546	3,916	2,070	1,846	1,289	893	1,177	396	1,450	2,627
6/23/2011	9,667	8,972	6,137	2,836	6,866	5,346	791	1,520	1,316	2,107
6/27/2011	6,146	7,415	4,947	2,468	5,184	4,114	832	1,069	1,399	2,231
6/29/2011	2,974									

Appendix 12. Control Reactor Biogas Production Data

Day	Date	Biogas Production Rate l / day	Accumulated Biogas Volume l
0	10/27/2010	2.63	2.08
1	10/28/2010	5.18	8.97
2	10/29/2010		8.97
5	11/1/2010	1.10	12.22
5	11/1/2010		12.22
5	11/1/2010		12.22
6	11/2/2010	3.39	15.86
7	11/3/2010	3.22	18.73
7	11/3/2010	4.93	20.27
8	11/4/2010	2.21	22.30
9	11/5/2010	3.01	25.14
12	11/8/2010	1.77	30.28
13	11/9/2010	6.05	37.49
14	11/10/2010	2.24	39.10
15	11/11/2010	3.81	44.07
16	11/12/2010	4.40	47.19
19	11/15/2010	8.79	73.54
20	11/16/2010	8.49	84.15
21	11/17/2010	3.34	86.67
22	11/18/2010	0.00	86.67
23	11/19/2010	6.40	91.18
26	11/22/2010	0.10	91.50
27	11/23/2010	0.00	91.50
30	11/26/2010	0.00	91.50
33	11/29/2010	0.00	91.50
34	11/30/2010	0.00	91.50
35	12/1/2010	0.39	91.78
36	12/2/2010	5.04	97.59
37	12/3/2010	1.03	98.46
40	12/6/2010	1.27	102.28
41	12/7/2010	2.90	105.99
42	12/8/2010	1.37	107.00
43	12/9/2010	4.78	113.09
44	12/10/2010	1.24	113.97
47	12/13/2010	1.60	118.76
48	12/14/2010	2.15	121.46
49	12/15/2010	2.75	123.52
50	12/16/2010	6.02	131.36
51	12/17/2010	6.17	136.33

Appendix 12. Control Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate l / day	Accumulated Biogas Volume l
54	12/20/2010	3.15	146.76
55	12/21/2010	0.19	146.94
57	12/23/2010	0.00	146.94
60	12/26/2010	0.00	146.94
63	12/29/2010	0.00	146.94
66	1/1/2011	0.00	146.94
69	1/4/2011	0.00	146.94
71	1/6/2011	0.02	146.97
72	1/7/2011	0.00	146.97
75	1/10/2011	12.80	186.70
77	1/12/2011	6.21	199.93
78	1/13/2011	6.32	204.83
79	1/14/2011	7.41	214.03
82	1/16/2011	3.90	222.82
83	1/18/2011	1.62	225.23
86	1/21/2011	1.02	228.31
89	1/24/2011	0.41	229.54
90	1/25/2011	0.70	230.24
92	1/27/2011	3.28	236.82
96	1/31/2011	1.92	244.52
100	2/4/2011	0.28	245.64
103	2/7/2011	0.00	245.64
104	2/8/2011	0.14	245.78
107	2/11/2011	0.87	248.40
110	2/14/2011	0.53	249.98
111	2/15/2011	12.00	292.85
112	2/16/2011	0.22	293.06
113	2/17/2011	1.65	294.71
114	2/18/2011	0.10	294.81
117	2/21/2011	8.79	321.34
118	2/22/2011	3.32	324.63
120	2/24/2011	9.85	344.30
121	2/25/2011	3.48	347.84
124	2/28/2011	7.72	371.04
125	3/1/2011	3.07	374.05
127	3/3/2011	7.34	388.68
128	3/4/2011	8.99	397.71
131	3/7/2011	5.08	412.97
132	3/8/2011	5.64	418.61
134	3/10/2011	2.74	424.07
135	3/11/2011	4.51	428.69

Appendix 12. Control Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate	Accumulated Biogas Volume
		l / day	l
139	3/15/2011	2.91	449.65
145	3/21/2011	2.29	463.34
146	3/22/2011	0.00	463.34
149	3/25/2011	6.05	481.54
152	3/28/2011	0.00	481.54
153	3/29/2011	8.24	489.55
155	3/31/2011	4.17	497.99
156	4/1/2011	1.94	499.91
159	4/4/2011	3.87	511.50
162	4/7/2011	0.00	511.50
167	4/12/2011	2.05	521.75
169	4/14/2011	2.24	526.23
173	4/18/2011	1.66	532.92
174	4/19/2011	3.69	536.49
176	4/21/2011	1.15	538.80
181	4/26/2011	2.39	550.73
183	4/28/2011	2.71	556.16
184	4/29/2011	1.07	557.52
187	5/2/2011	4.05	569.70
188	5/3/2011	3.90	572.54
190	5/5/2011	4.07	580.69
191	5/6/2011	12.94	593.99
194	5/9/2011	4.17	606.42
195	5/10/2011	1.23	607.64
196	5/11/2011	0.00	607.64
197	5/12/2011	2.09	609.74
201	5/16/2011	4.04	625.98
202	5/17/2011	4.71	630.64
203	5/18/2011	4.88	635.50
204	5/19/2011	4.78	640.33
205	5/20/2011	0.35	640.68
208	5/23/2011	6.35	659.79
209	5/24/2011	1.60	661.44
210	5/25/2011	4.67	666.09
211	5/26/2011	3.66	669.59
216	5/31/2011	5.21	695.70
217	6/1/2011	1.48	697.10
218	6/2/2011	3.80	701.13
219	6/3/2011	18.16	719.13
222	6/6/2011	6.72	739.13
225	6/9/2011	3.45	749.63

Appendix 12. Control Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate l / day	Accumulated Biogas Volume l
226	6/10/2011	6.11	755.63
229	6/13/2011	2.64	763.50
230	6/14/2011	2.38	765.88
231	6/15/2011	3.36	769.38
232	6/16/2011	3.11	772.38
233	6/17/2011	4.55	776.88
236	6/20/2011	1.68	781.96
237	6/21/2011	1.99	783.95
238	6/22/2011	2.07	785.98
239	6/23/2011	3.00	787.98
240	6/24/2011	3.50	792.98
243	6/27/2011	4.00	796.97
244	6/28/2011	3.49	800.47
245	6/29/2011	4.45	804.87
246	6/30/2011	5.01	809.87
247	7/1/2011	2.56	812.47
254	7/8/2011	0.35	814.92

Appendix 13. Glycerol treatment reactor glycerol inclusion rate

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
	10/27/2010	1.00	13.00	0.36%
1	10/28/2010	1.00	13.00	0.36%
2	10/29/2010	1.00	13.00	0.36%
3	10/30/2010	1.00	13.00	0.36%
4	10/31/2010	1.00	13.00	0.36%
5	11/1/2010	1.00	13.00	0.36%
6	11/2/2010	1.00	13.00	0.36%
7	11/3/2010	1.25	25.99	0.45%
8	11/4/2010	1.50	25.99	0.54%
9	11/5/2010	1.75	25.99	0.63%
10	11/6/2010	2.00	25.99	0.72%
11	11/7/2010	2.00	25.99	0.72%
12	11/8/2010	2.00	25.99	0.72%
13	11/9/2010	2.00	25.99	0.72%
14	11/10/2010	2.25	38.99	0.81%
15	11/11/2010	2.50	38.99	0.90%
16	11/12/2010	2.75	38.99	0.99%
17	11/13/2010	3.00	38.99	1.08%
18	11/14/2010	3.00	38.99	1.08%
19	11/15/2010	3.00	38.99	1.08%
20	11/16/2010	3.00	38.99	1.08%
21	11/17/2010	3.12	45.00	1.12%
22	11/18/2010	3.23	45.00	1.17%
23	11/19/2010	3.35	45.00	1.21%
24	11/20/2010	3.46	45.00	1.25%
25	11/21/2010	3.46	45.00	1.25%
26	11/22/2010	3.46	45.00	1.25%
27	11/23/2010	3.46	45.00	1.25%
28	11/24/2010	3.46	45.00	1.25%
29	11/25/2010	3.46	45.00	1.25%
30	11/26/2010	3.46	45.00	1.25%
31	11/27/2010	3.46	45.00	1.25%
32	11/28/2010	3.46	45.00	1.25%
33	11/29/2010	3.46	45.00	1.25%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
34	11/30/2010	3.46	45.00	1.25%
35	12/1/2010	3.35	38.99	1.21%
36	12/2/2010	3.23	38.99	1.17%
37	12/3/2010	3.12	38.99	1.12%
38	12/4/2010	3.00	38.99	1.08%
39	12/5/2010	3.00	38.99	1.08%
40	12/6/2010	3.00	38.99	1.08%
41	12/7/2010	3.00	38.99	1.08%
42	12/8/2010	3.06	41.98	1.10%
43	12/9/2010	3.12	41.98	1.12%
44	12/10/2010	3.17	41.98	1.15%
45	12/11/2010	3.23	41.98	1.17%
46	12/12/2010	3.23	41.98	1.17%
47	12/13/2010	3.23	41.98	1.17%
48	12/14/2010	3.23	41.98	1.17%
49	12/15/2010	3.23	41.98	1.17%
50	12/16/2010	3.23	41.98	1.17%
51	12/17/2010	3.23	41.98	1.17%
52	12/18/2010	3.23	41.98	1.17%
53	12/19/2010	3.23	41.98	1.17%
54	12/20/2010	3.23	41.98	1.17%
55	12/21/2010	0.00	0.00	0.00%
56	12/22/2010	0.00	0.00	0.00%
57	12/23/2010	0.00	0.00	0.00%
58	12/24/2010	0.00	0.00	0.00%
59	12/25/2010	0.00	0.00	0.00%
60	12/26/2010	0.00	0.00	0.00%
61	12/27/2010	0.00	0.00	0.00%
62	12/28/2010	0.00	0.00	0.00%
63	12/29/2010	0.00	0.00	0.00%
64	12/30/2010	0.00	0.00	0.00%
65	12/31/2010	0.00	0.00	0.00%
66	1/1/2011	0.00	0.00	0.00%
67	1/2/2011	0.00	0.00	0.00%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
68	1/3/2011	0.00	0.00	0.00%
69	1/4/2011	0.00	0.00	0.00%
70	1/5/2011	0.00	0.00	0.00%
71	1/6/2011	3.00	38.99	1.08%
72	1/7/2011	3.00	38.99	1.08%
73	1/8/2011	3.00	38.99	1.08%
74	1/9/2011	3.00	38.99	1.08%
75	1/10/2011	3.00	38.99	1.08%
76	1/11/2011	3.00	38.99	1.08%
77	1/12/2011	3.00	38.99	1.08%
78	1/13/2011	3.00	38.99	1.08%
79	1/14/2011	3.00	38.99	1.08%
80	1/15/2011	3.00	38.99	1.08%
81	1/16/2011	3.00	38.99	1.08%
82	1/17/2011	3.00	38.99	1.08%
83	1/18/2011	3.00	38.99	1.08%
84	1/19/2011	3.00	38.99	1.08%
85	1/20/2011	3.00	38.99	1.08%
86	1/21/2011	3.00	38.99	1.08%
87	1/22/2011	3.00	38.99	1.08%
88	1/23/2011	3.00	38.99	1.08%
89	1/24/2011	3.00	38.99	1.08%
90	1/25/2011	3.00	38.99	1.08%
91	1/26/2011	3.10	44.00	1.12%
92	1/27/2011	3.19	44.00	1.15%
93	1/28/2011	3.29	44.00	1.19%
94	1/29/2011	3.39	44.00	1.22%
95	1/30/2011	3.39	44.00	1.22%
96	1/31/2011	3.39	44.00	1.22%
97	2/1/2011	3.39	44.00	1.22%
98	2/2/2011	3.39	44.00	1.22%
99	2/3/2011	3.39	44.00	1.22%
100	2/4/2011	3.39	44.00	1.22%
101	2/5/2011	3.39	44.00	1.22%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate g COD / l / day	Glycerol Treatment Glycerol Daily Volume added to Tank ml/day	Glycerol Treatment Inclusion Rate %
102	2/6/2011	3.39	44.00	1.22%
103	2/7/2011	3.39	44.00	1.22%
104	2/8/2011	3.39	44.00	1.22%
105	2/9/2011	3.39	44.00	1.22%
106	2/10/2011	3.39	44.00	1.22%
107	2/11/2011	3.39	44.00	1.22%
108	2/12/2011	3.39	44.00	1.22%
109	2/13/2011	3.39	44.00	1.22%
110	2/14/2011	3.39	44.00	1.22%
111	2/15/2011	3.39	44.00	1.22%
112	2/16/2011	3.23	36.00	1.17%
113	2/17/2011	3.08	36.00	1.11%
114	2/18/2011	2.92	36.00	1.06%
115	2/19/2011	2.77	36.00	1.00%
116	2/20/2011	2.77	36.00	1.00%
117	2/21/2011	2.77	36.00	1.00%
118	2/22/2011	2.77	36.00	1.00%
119	2/23/2011	2.77	36.00	1.00%
120	2/24/2011	2.77	36.00	1.00%
121	2/25/2011	2.77	36.00	1.00%
122	2/26/2011	2.08	0.00	1.00%
123	2/27/2011	2.08	36.00	1.00%
124	2/28/2011	1.89	26.00	0.93%
125	3/1/2011	1.19	0.00	0.68%
126	3/2/2011	1.19	0.00	0.43%
127	3/3/2011	0.50	0.00	0.18%
128	3/4/2011	0.25	13.00	0.09%
129	3/5/2011	0.50	13.00	0.18%
130	3/6/2011	0.75	13.00	0.27%
131	3/7/2011	1.00	13.00	0.36%
132	3/8/2011	1.00	13.00	0.36%
133	3/9/2011	1.00	13.00	0.36%
134	3/10/2011	1.00	13.00	0.36%
135	3/11/2011	1.00	13.00	0.36%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
136	3/12/2011	1.00	13.00	0.36%
137	3/13/2011	1.00	13.00	0.36%
138	3/14/2011	1.00	13.00	0.36%
139	3/15/2011	1.00	13.00	0.36%
140	3/16/2011	1.00	13.00	0.36%
141	3/17/2011	1.00	13.00	0.36%
142	3/18/2011	1.00	13.00	0.36%
143	3/19/2011	1.00	13.00	0.36%
144	3/20/2011	1.00	13.00	0.36%
145	3/21/2011	1.00	13.00	0.36%
146	3/22/2011	1.00	13.00	0.36%
147	3/23/2011	1.00	13.00	0.36%
148	3/24/2011	1.00	13.00	0.36%
149	3/25/2011	1.00	13.00	0.36%
150	3/26/2011	1.00	13.00	0.36%
151	3/27/2011	1.00	13.00	0.36%
152	3/28/2011	1.00	13.00	0.36%
153	3/29/2011	1.00	13.00	0.36%
154	3/30/2011	1.00	13.00	0.36%
155	3/31/2011	1.00	13.00	0.36%
156	4/1/2011	1.00	13.00	0.36%
157	4/2/2011	1.00	13.00	0.36%
158	4/3/2011	1.00	13.00	0.36%
159	4/4/2011	1.00	13.00	0.36%
160	4/5/2011	1.00	13.00	0.36%
161	4/6/2011	1.00	13.00	0.36%
162	4/7/2011	1.04	15.00	0.38%
163	4/8/2011	1.08	15.00	0.39%
164	4/9/2011	1.12	15.00	0.40%
165	4/10/2011	1.15	15.00	0.42%
166	4/11/2011	1.15	15.00	0.42%
167	4/12/2011	1.15	15.00	0.42%
168	4/13/2011	1.15	15.00	0.42%
169	4/14/2011	1.15	15.00	0.42%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
170	4/15/2011	1.15	15.00	0.42%
171	4/16/2011	1.15	15.00	0.42%
172	4/17/2011	1.15	15.00	0.42%
173	4/18/2011	1.15	15.00	0.42%
174	4/19/2011	1.15	15.00	0.42%
175	4/20/2011	1.21	18.00	0.44%
176	4/21/2011	1.27	18.00	0.46%
177	4/22/2011	1.33	18.00	0.48%
178	4/23/2011	1.39	18.00	0.50%
179	4/24/2011	1.39	18.00	0.50%
180	4/25/2011	1.39	18.00	0.50%
181	4/26/2011	1.39	18.00	0.50%
182	4/27/2011	1.46	22.00	0.53%
183	4/28/2011	1.54	22.00	0.56%
184	4/29/2011	1.62	22.00	0.58%
185	4/30/2011	1.69	22.00	0.61%
186	5/1/2011	1.69	22.00	0.61%
187	5/2/2011	1.69	22.00	0.61%
188	5/3/2011	1.69	22.00	0.61%
189	5/4/2011	1.77	26.00	0.64%
190	5/5/2011	1.85	26.00	0.67%
191	5/6/2011	1.92	26.00	0.69%
192	5/7/2011	2.00	26.00	0.72%
193	5/8/2011	2.00	26.00	0.72%
194	5/9/2011	2.00	26.00	0.72%
195	5/10/2011	2.00	26.00	0.72%
196	5/11/2011	2.10	31.00	0.76%
197	5/12/2011	2.19	31.00	0.79%
198	5/13/2011	2.29	31.00	0.83%
199	5/14/2011	2.39	31.00	0.86%
200	5/15/2011	2.39	31.00	0.86%
201	5/16/2011	2.39	31.00	0.86%
202	5/17/2011	2.39	31.00	0.86%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
203	5/18/2011	2.39	31.00	0.86%
204	5/19/2011	2.39	31.00	0.86%
205	5/20/2011	2.39	31.00	0.86%
206	5/21/2011	2.39	31.00	0.86%
207	5/22/2011	2.39	31.00	0.86%
208	5/23/2011	2.39	31.00	0.86%
209	5/24/2011	2.39	31.00	0.86%
210	5/25/2011	2.39	31.00	0.86%
211	5/26/2011	2.39	31.00	0.86%
212	5/27/2011	2.42	33.00	0.88%
213	5/28/2011	2.46	33.00	0.89%
214	5/29/2011	2.50	33.00	0.90%
215	5/30/2011	2.54	33.00	0.92%
216	5/31/2011	2.54	33.00	0.92%
217	6/1/2011	2.54	33.00	0.92%
218	6/2/2011	2.54	33.00	0.92%
219	6/3/2011	2.54	33.00	0.92%
220	6/4/2011	2.54	33.00	0.92%
221	6/5/2011	2.54	33.00	0.92%
222	6/6/2011	2.54	33.00	0.92%
223	6/7/2011	2.54	33.00	0.92%
224	6/8/2011	2.54	33.00	0.92%
225	6/9/2011	2.54	33.00	0.92%
226	6/10/2011	2.54	33.00	0.92%
227	6/11/2011	2.54	33.00	0.92%
228	6/12/2011	2.54	33.00	0.92%
229	6/13/2011	2.54	33.00	0.92%
230	6/14/2011	2.54	33.00	0.92%
231	6/15/2011	2.60	36.00	0.94%
232	6/16/2011	2.65	36.00	0.96%
233	6/17/2011	2.71	36.00	0.98%
234	6/18/2011	2.77	36.00	1.00%
235	6/19/2011	2.77	36.00	1.00%
236	6/20/2011	2.77	36.00	1.00%

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate	Glycerol Treatment Glycerol Daily Volume added to Tank	Glycerol Treatment Inclusion Rate
		g COD / l / day	ml/day	%
237	6/21/2011	2.77	36.00	1.00
238	6/22/2011	2.77	36.00	1.00
239	6/23/2011	2.77	36.00	1.00
240	6/24/2011	2.77	36.00	1.00
241	6/25/2011	2.77	36.00	1.00
242	6/26/2011	2.77	36.00	1.00
243	6/27/2011	2.77	36.00	1.00
244	6/28/2011	2.77	36.00	1.00
245	6/29/2011	2.77	36.00	1.00
246	6/30/2011	2.77	36.00	1.00
247	7/1/2011	2.77	36.00	1.00
248	7/2/2011	2.77	36.00	1.00
249	7/3/2011	2.77	36.00	1.00
250	7/4/2011	2.77	36.00	1.00
251	7/5/2011	2.77	36.00	1.00
252	7/6/2011	2.77	36.00	1.00
253	7/7/2011	2.77	36.00	1.00
254	7/8/2011	2.77	36.00	1.00
255	7/9/2011	2.77	36.00	1.00
256	7/10/2011	2.77	36.00	1.00
257	7/11/2011	2.77	36.00	1.00
258	7/12/2011	2.77	36.00	1.00
259	7/13/2011	2.77	36.00	1.00
260	7/14/2011	2.77	36.00	1.00
261	7/15/2011	2.77	36.00	1.00
262	7/16/2011	2.77	36.00	1.00
263	7/17/2011	2.77	36.00	1.00
264	7/18/2011	2.77	36.00	1.00
265	7/19/2011	2.77	36.00	1.00
266	7/20/2011	2.77	36.00	1.00
267	7/21/2011	2.77	36.00	1.00
268	7/22/2011	2.77	36.00	1.00
269	7/23/2011	2.77	36.00	1.00
270	7/24/2011	2.77	36.00	1.00

Appendix 13. Glycerol treatment reactor glycerol inclusion rate continued

Day	Date	Glycerol Treatment Glycerol Loading Rate g COD / l / day	Glycerol Treatment Glycerol Daily Volume added to Tank ml/day	Glycerol Treatment Inclusion Rate %
273	7/27/2011	2.77	36.00	1.00%
274	7/28/2011	2.77	36.00	1.00%
275	7/29/2011	2.77	36.00	1.00%
276	7/30/2011	2.77	36.00	1.00%
277	7/31/2011	2.77	36.00	1.00%
278	8/1/2011	2.77	36.00	1.00%
279	8/2/2011	2.77	36.00	1.00%
280	8/3/2011	2.77	36.00	1.00%
281	8/4/2011	2.77	36.00	1.00%
282	8/5/2011	2.77	36.00	1.00%
283	8/6/2011	2.77	36.00	1.00%
284	8/7/2011	2.77	36.00	1.00%
285	8/8/2011	2.77	36.00	1.00%
286	8/9/2011	2.77	36.00	1.00%
287	8/10/2011	2.77	36.00	1.00%
288	8/11/2011	2.77	36.00	1.00%
289	8/12/2011	2.77	36.00	1.00%
290	8/13/2011	2.77	36.00	1.00%
291	8/14/2011	2.77	36.00	1.00%
292	8/15/2011	2.77	36.00	1.00%
293	8/16/2011	2.77	36.00	1.00%
294	8/17/2011	2.77	36.00	1.00%
295	8/18/2011	2.77	36.00	1.00%
296	8/19/2011	2.77	36.00	1.00%
297	8/20/2011	2.77	36.00	1.00%

Appendix 14. Glycerol Treatment Influent Parameters

Date	COD mg/l	TS mg/l	VS mg/l	TFS mg/l	TSS mg/l	VSS mg/l	VDS mg/l	FSS mg/l	FDS mg/l	TDS mg/l
10/15/10	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/21/10	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/29/10	13,314	7,179	5,396	1,783	5,133	4,249	1,147	884	899	2,046
11/3/10	11,676	7,358	5,594	1,764	5,234	4,503	1,091	731	1,033	2,124
11/11/10	24,020	27,071	22,622	4,449	24,218	21,183	1,439	3,035	1,414	2,853
11/17/10	21,959	13,496	10,509	2,987	8,817	7,555	2,954	1,262	1,725	4,679
11/29/10	19,926	9,502	7,107	2,395	5,912	5,080	2,027	832	1,563	3,590
12/1/10	17,789	6,322	4,389	1,933	2,156	1,800	2,589	356	1,577	4,166
12/6/10	19,107	9,782	7,260	2,522	4,096	3,322	3,938	774	1,748	5,686
12/10/10	21,564	14,733	11,591	3,142	9,067	7,828	3,763	1,239	1,903	5,666
12/13/10	22,505	17,197	13,903	3,294	11,459	9,879	4,024	1,580	1,714	5,738
12/15/10	20,001	8,630	5,843	2,787	4,182	3,531	2,312	651	2,136	4,448
1/13/11	21,903	17,779	14,341	3,438	14,210	12,373	1,968	1,837	1,601	3,569
1/18/11	19,700	8,539	6,266	2,273	5,113	4,276	1,990	837	1,436	3,426
1/20/11	20,933	11,961	9,488	2,473	8,108	6,948	2,540	1,160	1,313	3,853
1/25/11	19,785	8,194	6,183	2,011	4,502	3,832	2,351	670	1,341	3,692
1/27/11	20,660	13,997	11,309	2,688	11,379	9,955	1,354	1,424	1,264	2,618
2/8/11	21,272	16,966	14,316	2,650	12,909	11,795	2,521	1,114	1,536	4,057
2/15/11	26,129	25,842	22,348	3,494	21,270	19,052	3,296	2,218	1,276	4,572
2/17/11	20,020	9,599	7,288	2,311	4,833	4,156	3,132	677	1,634	4,766
2/22/11	20,820	14,012	11,543	2,469	11,419	10,149	1,394	1,270	1,199	2,593

Appendix 14. Glycerol Treatment Influent Parameters continued

Date	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
2/24/11	20,368	10,128	8,059	2,069	6,929	6,209	1,850	720	1,349	3,199
3/1/11	19,888	10,163	8,144	2,019	7,589	6,776	1,368	813	1,206	2,574
3/3/11	19,747	10,616	8,587	2,029	7,578	6,708	1,879	870	1,159	3,038
3/8/11	16,914	9,396	7,540	1,856	7,027	6,203	1,337	824	1,032	2,369
3/10/11	19,239	11,759	9,548	2,211	9,451	8,267	1,281	1,184	1,027	2,308
3/15/11	17,328	10,140	8,102	2,038	8,000	7,015	1,087	985	1,053	2,140
3/17/11	15,653	8,486	6,693	1,793	6,324	5,565	1,128	759	1,034	2,162
3/22/11	19,935	15,624	13,057	2,567	13,447	11,963	1,094	1,484	1,083	2,177
3/29/11	17,338	10,893	8,902	1,991	8,743	7,751	1,151	992	999	2,150
3/31/11	17,469	10,599	8,504	2,095	8,228	7,113	1,391	1,115	980	2,371
4/7/11	17,196	12,001	9,988	2,013	9,505	8,442	1,546	1,063	950	2,496
4/12/11	11,737	6,256	4,943	1,313	4,200	3,676	1,267	524	789	2,056
4/14/11	10,580	5,866	4,539	1,327	3,526	3,085	1,454	441	886	2,340
4/19/11	11,464	5,241	3,651	1,590	3,157	2,687	964	470	1,120	2,084
4/21/11	16,359	8,302	6,407	1,895	6,149	5,267	1,140	882	1,013	2,153
4/26/11	13,935	6,524	4,748	1,776	4,153	3,398	1,350	755	1,021	2,371
5/3/11	14,589	7,644	5,712	1,932	4,087	3,435	2,277	652	1,280	3,557
5/5/11	23,578	11,257	8,739	2,518	8,811	7,536	1,203	1,275	1,243	2,446
5/10/11	29,296	22,733	19,039	3,694	19,436	17,230	1,809	2,206	1,488	3,297
5/17/11	40,026	26,067	22,127	3,940	22,544	20,168	1,959	2,376	1,564	3,523
5/19/11	29,672	16,876	13,962	2,913	12,494	10,870	3,092	1,624	1,289	4,381
5/25/11	20,495	11,293	8,668	2,625	9,091	7,898	770	1,192	1,432	2,202
5/27/11		8,026	6,257	1,769	6,631	5,744	512	887	882	1,394

Appendix 14. Glycerol Treatment Influent Parameters continued

Date	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
5/31/11		15,651	12,358	3,293	12,212	10,054	2,303	2,158	1,136	3,439
6/7/11	15,248	7,598	5,254	2,343	4,587	3,618	1,636	969	1,375	3,011
6/8/11	16,683	9,833	7,370	2,463	6,399	5,465	1,905	934	1,529	3,434
6/13/11	17,907	9,666	7,326	2,340	5,351	4,737	2,589	614	1,726	4,314
6/16/11	24,543	14,907	12,061	2,846	11,788	10,494	1,567	1,293	1,552	3,119
6/20/11	31,720	9,657	7,054	2,602	4,577	3,666	3,389	911	1,691	5,080
6/27/11	34,214	17,989	14,733	3,256	13,921	12,260	2,473	1,661	1,594	4,068
6/29/11	40,214									

Appendix 15. Glycerol Treatment Reactor Effluent Parameters

Date	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
10/15/2010	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/21/2010	8,659	8,093	5,723	2,370	5,517	4,603	1,120	914	1,456	2,576
10/29/2010	13,314	7,179	5,396	1,783	5,133	4,249	1,147	884	899	2,046
11/3/2010	11,676	7,358	5,594	1,764	5,234	4,503	1,091	731	1,033	2,124
11/11/2010	24,020	27,071	22,622	4,449	24,218	21,183	1,439	3,035	1,414	2,853
11/17/2010	21,959	13,496	10,509	2,987	8,817	7,555	2,954	1,262	1,725	4,679
11/29/2010	19,926	9,502	7,107	2,395	5,912	5,080	2,027	832	1,563	3,590
12/1/2010	17,789	6,322	4,389	1,933	2,156	1,800	2,589	356	1,577	4,166
12/6/2010	19,107	9,782	7,260	2,522	4,096	3,322	3,938	774	1,748	5,686
12/10/2010	21,564	14,733	11,591	3,142	9,067	7,828	3,763	1,239	1,903	5,666
12/13/2010	22,505	17,197	13,903	3,294	11,459	9,879	4,024	1,580	1,714	5,738
12/15/2010	20,001	8,630	5,843	2,787	4,182	3,531	2,312	651	2,136	4,448
1/13/2011	21,903	17,779	14,341	3,438	14,210	12,373	1,968	1,837	1,601	3,569
1/18/2011	19,700	8,539	6,266	2,273	5,113	4,276	1,990	837	1,436	3,426
1/20/2011	20,933	11,961	9,488	2,473	8,108	6,948	2,540	1,160	1,313	3,853
1/25/2011	19,785	8,194	6,183	2,011	4,502	3,832	2,351	670	1,341	3,692
1/27/2011	20,660	13,997	11,309	2,688	11,379	9,955	1,354	1,424	1,264	2,618
2/8/2011	21,272	16,966	14,316	2,650	12,909	11,795	2,521	1,114	1,536	4,057
2/15/2011	26,129	25,842	22,348	3,494	21,270	19,052	3,296	2,218	1,276	4,572
2/17/2011	20,020	9,599	7,288	2,311	4,833	4,156	3,132	677	1,634	4,766
2/22/2011	20,820	14,012	11,543	2,469	11,419	10,149	1,394	1,270	1,199	2,593
2/24/2011	20,368	10,128	8,059	2,069	6,929	6,209	1,850	720	1,349	3,199
3/1/2011	19,888	10,163	8,144	2,019	7,589	6,776	1,368	813	1,206	2,574

Appendix 15. Glycerol Treatment Reactor Effluent Parameters continued

Date	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
3/3/2011	19,747	10,616	8,587	2,029	7,578	6,708	1,879	870	1,159	3,038
3/8/2011	16,914	9,396	7,540	1,856	7,027	6,203	1,337	824	1,032	2,369
3/10/2011	19,239	11,759	9,548	2,211	9,451	8,267	1,281	1,184	1,027	2,308
3/15/2011	17,328	10,140	8,102	2,038	8,000	7,015	1,087	985	1,053	2,140
3/17/2011	15,653	8,486	6,693	1,793	6,324	5,565	1,128	759	1,034	2,162
3/22/2011	19,935	15,624	13,057	2,567	13,447	11,963	1,094	1,484	1,083	2,177
3/29/2011	17,338	10,893	8,902	1,991	8,743	7,751	1,151	992	999	2,150
3/31/2011	17,469	10,599	8,504	2,095	8,228	7,113	1,391	1,115	980	2,371
4/7/2011	17,196	12,001	9,988	2,013	9,505	8,442	1,546	1,063	950	2,496
4/12/2011	11,737	6,256	4,943	1,313	4,200	3,676	1,267	524	789	2,056
4/14/2011	10,580	5,866	4,539	1,327	3,526	3,085	1,454	441	886	2,340
4/19/2011	11,464	5,241	3,651	1,590	3,157	2,687	964	470	1,120	2,084
4/21/2011	16,359	8,302	6,407	1,895	6,149	5,267	1,140	882	1,013	2,153
4/26/2011	13,935	6,524	4,748	1,776	4,153	3,398	1,350	755	1,021	2,371
5/3/2011	14,589	7,644	5,712	1,932	4,087	3,435	2,277	652	1,280	3,557
5/5/2011	23,578	11,257	8,739	2,518	8,811	7,536	1,203	1,275	1,243	2,446
5/10/2011	29,296	22,733	19,039	3,694	19,436	17,230	1,809	2,206	1,488	3,297
5/17/2011	40,026	26,067	22,127	3,940	22,544	20,168	1,959	2,376	1,564	3,523
5/19/2011	29,672	16,876	13,962	2,913	12,494	10,870	3,092	1,624	1,289	4,381
5/25/2011	20,495	11,293	8,668	2,625	9,091	7,898	770	1,192	1,432	2,202
5/27/2011		8,026	6,257	1,769	6,631	5,744	512	887	882	1,394
5/31/2011		15,651	12,358	3,293	12,212	10,054	2,303	2,158	1,136	3,439
6/7/2011	15,248	7,598	5,254	2,343	4,587	3,618	1,636	969	1,375	3,011

Appendix 15. Glycerol Treatment Reactor Effluent Parameters continued

Date	COD	TS	VS	TFS	TSS	VSS	VDS	FSS	FDS	TDS
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
6/8/2011	16,683	9,833	7,370	2,463	6,399	5,465	1,905	934	1,529	3,434
6/13/2011	17,907	9,666	7,326	2,340	5,351	4,737	2,589	614	1,726	4,314
6/16/2011	24,543	14,907	12,061	2,846	11,788	10,494	1,567	1,293	1,552	3,119
6/20/2011	31,720	9,657	7,054	2,602	4,577	3,666	3,389	911	1,691	5,080
6/27/2011	34,214	17,989	14,733	3,256	13,921	12,260	2,473	1,661	1,594	4,068
6/29/2011	40,214									

Appendix 16. Glycerol Treatment Reactor Biogas Production Data

Day	Date	Biogas Production Rate l/day	Accumulated Biogas Volume l
0	10/27/2010	6.53	6.53
1	10/28/2010	7.76	14.29
2	10/29/2010	8.69	22.99
3	10/30/2010	9.61	32.60
4	10/31/2010	8.74	41.34
5	11/1/2010	8.74	50.08
6	11/2/2010	8.74	58.82
7	11/3/2010	8.82	67.64
8	11/4/2010	10.46	78.10
9	11/5/2010	12.28	90.39
10	11/6/2010	12.77	103.16
11	11/7/2010	12.77	115.93
12	11/8/2010	12.77	128.69
13	11/9/2010	12.73	141.42
14	11/10/2010	14.37	155.79
15	11/11/2010	16.47	172.26
16	11/12/2010	16.89	189.15
17	11/13/2010	17.79	206.94
18	11/14/2010	17.79	224.73
19	11/15/2010	17.79	242.53
20	11/16/2010	17.96	260.49
21	11/17/2010	17.52	278.00
22	11/18/2010	18.75	296.76
23	11/19/2010	18.72	315.48
24	11/20/2010	19.09	334.57
25	11/21/2010	19.09	353.66
26	11/22/2010	19.09	372.76
27	11/23/2010	16.33	389.08
28	11/24/2010	14.98	404.06
29	11/25/2010	14.98	419.04
30	11/26/2010	14.98	434.02
31	11/27/2010	14.20	448.22
32	11/28/2010	14.20	462.42
33	11/29/2010	14.20	476.62
34	11/30/2010	12.54	489.16
35	12/1/2010	13.71	502.87
36	12/2/2010	12.89	515.76
37	12/3/2010	12.58	528.34
38	12/4/2010	12.49	540.83
39	12/5/2010	12.49	553.32
40	12/6/2010	12.49	565.80
41	12/7/2010	12.98	578.78

Appendix 16. Glycerol Treatment Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate	Accumulated Biogas Volume
		l/day	l
42	12/8/2010	13.35	592.13
43	12/9/2010	13.16	605.28
44	12/10/2010	13.44	618.72
45	12/11/2010	13.17	631.89
46	12/12/2010	13.17	645.06
47	12/13/2010	13.17	658.23
48	12/14/2010	15.69	673.92
49	12/15/2010	18.19	692.11
50	12/16/2010	18.19	710.30
51	12/17/2010	19.24	729.54
52	12/18/2010	19.24	748.78
53	12/19/2010	19.24	768.02
54	12/20/2010	19.24	787.26
55	12/21/2010	20.67	807.93
56	12/22/2010	15.89	823.82
57	12/23/2010	15.21	839.03
58	12/24/2010	13.91	852.94
59	12/25/2010	11.83	864.77
60	12/26/2010	9.36	874.13
61	12/27/2010	6.11	880.24
62	12/28/2010	5.98	886.22
63	12/29/2010	6.37	892.59
64	12/30/2010	5.46	898.05
65	12/31/2010	3.90	901.95
66	1/1/2011	3.51	905.46
67	1/2/2011	3.64	909.10
68	1/3/2011	2.99	912.09
69	1/4/2011	2.34	914.43
70	1/5/2011	0.00	914.43
71	1/6/2011	7.02	921.45
72	1/7/2011	12.61	934.06
73	1/8/2011	17.29	951.35
74	1/9/2011	18.46	969.81
75	1/10/2011	15.73	985.54
76	1/11/2011	14.69	1000.23
77	1/12/2011	15.99	1016.22
78	1/13/2011	17.68	1033.90
79	1/14/2011	18.07	1051.97
80	1/15/2011	18.46	1070.43
81	1/16/2011	18.85	1089.28
82	1/17/2011	20.28	1109.56
83	1/18/2011	18.07	1127.63

Appendix 16. Glycerol Treatment Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate l/day	Accumulated Biogas Volume l
84	1/19/2011	17.42	1145.05
85	1/20/2011	17.94	1162.99
86	1/21/2011	18.98	1181.97
87	1/22/2011	20.93	1202.90
88	1/23/2011	22.88	1225.78
89	1/24/2011	24.83	1250.61
90	1/25/2011	26.78	1277.39
91	1/26/2011	26.13	1303.52
92	1/27/2011	26.52	1330.04
93	1/28/2011	26.13	1356.17
94	1/29/2011	27.04	1383.21
95	1/30/2011	28.99	1412.20
96	1/31/2011	27.82	1440.02
97	2/1/2011	23.27	1463.29
98	2/2/2011	21.84	1485.13
99	2/3/2011	20.54	1505.67
100	2/4/2011	18.33	1524.00
101	2/5/2011	30.03	1554.03
102	2/6/2011	28.21	1582.24
103	2/7/2011	26.39	1608.63
104	2/8/2011	23.14	1631.77
105	2/9/2011	20.67	1652.44
106	2/10/2011	19.76	1672.20
107	2/11/2011	18.33	1690.53
108	2/12/2011	20.80	1711.33
109	2/13/2011	20.54	1731.87
110	2/14/2011	13.52	1745.39
111	2/15/2011	13.00	1758.39
112	2/16/2011	13.78	1772.17
113	2/17/2011	13.78	1785.95
114	2/18/2011	11.96	1797.91
115	2/19/2011	11.05	1808.96
116	2/20/2011	10.01	1818.97
117	2/21/2011	5.59	1824.56
118	2/22/2011	5.85	1830.41
119	2/23/2011	9.36	1839.77
120	2/24/2011	5.98	1845.75
121	2/25/2011	5.59	1851.34
122	2/26/2011	4.59	1855.93
123	2/27/2011	4.94	1860.87
124	2/28/2011	3.77	1864.64
125	3/1/2011	17.03	1881.67

Appendix 16. Glycerol Treatment Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate l/day	Accumulated Biogas Volume l
126	3/2/2011	16.90	1898.57
127	3/3/2011	17.16	1915.73
128	3/4/2011	15.08	1930.81
129	3/5/2011	14.30	1945.11
130	3/6/2011	14.69	1959.80
131	3/7/2011	10.27	1970.07
132	3/8/2011	8.84	1978.91
133	3/9/2011	6.37	1985.28
134	3/10/2011	3.77	1989.05
135	3/11/2011	4.68	1993.73
136	3/12/2011	3.38	1997.11
137	3/13/2011	1.69	1998.80
138	3/14/2011	3.21	2002.01
139	3/15/2011	2.64	2004.65
140	3/16/2011	2.21	2006.86
141	3/17/2011	2.08	2008.94
142	3/18/2011	1.43	2010.37
143	3/19/2011	2.73	2013.10
144	3/20/2011	2.73	2015.83
145	3/21/2011	5.72	2021.55
146	3/22/2011	8.84	2030.39
147	3/23/2011	8.19	2038.58
148	3/24/2011	7.93	2046.51
149	3/25/2011	8.97	2055.48
150	3/26/2011	4.68	2060.16
151	3/27/2011	2.21	2062.37
152	3/28/2011	4.42	2066.79
153	3/29/2011	1.95	2068.74
154	3/30/2011	1.43	2070.17
155	3/31/2011	0.13	2070.30
156	4/1/2011	1.56	2071.86
157	4/2/2011	6.89	2078.75
158	4/3/2011	7.28	2086.03
159	4/4/2011	6.76	2092.79
160	4/5/2011	7.54	2100.33
161	4/6/2011	5.72	2106.05
162	4/7/2011	8.58	2114.63
163	4/8/2011	8.58	2123.21
164	4/9/2011	7.15	2130.36
165	4/10/2011	4.29	2134.65
166	4/11/2011	2.60	2137.25
167	4/12/2011	3.12	2140.37

Appendix 16. Glycerol Treatment Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate l/day	Accumulated Biogas Volume l
168	4/13/2011	2.21	2142.58
169	4/14/2011	3.64	2146.22
170	4/15/2011	0.91	2147.13
171	4/16/2011	0.91	2148.04
172	4/17/2011	1.82	2149.86
173	4/18/2011	1.95	2151.81
174	4/19/2011	8.84	2160.65
175	4/20/2011	9.75	2170.40
176	4/21/2011	9.75	2180.15
177	4/22/2011	7.93	2188.08
178	4/23/2011	6.37	2194.45
179	4/24/2011	4.29	2198.74
180	4/25/2011	6.37	2205.11
181	4/26/2011	6.89	2212.00
182	4/27/2011	9.75	2221.75
183	4/28/2011	12.87	2234.62
184	4/29/2011	14.43	2249.05
185	4/30/2011	13.65	2262.70
186	5/1/2011	11.05	2273.75
187	5/2/2011	10.66	2284.41
188	5/3/2011	12.22	2296.63
189	5/4/2011	14.95	2311.58
190	5/5/2011	14.69	2326.27
191	5/6/2011	13.78	2340.05
192	5/7/2011	13.13	2353.18
193	5/8/2011	13.13	2366.31
194	5/9/2011	14.56	2380.87
195	5/10/2011	14.17	2395.04
196	5/11/2011	14.43	2409.47
197	5/12/2011	10.27	2419.74
198	5/13/2011	10.92	2430.66
201	5/16/2011	12.74	2443.40
202	5/17/2011	22.10	2465.50
203	5/18/2011	24.44	2489.94
204	5/19/2011	25.35	2515.29
205	5/20/2011	26.26	2541.55
206	5/21/2011	26.52	2568.07
207	5/22/2011	24.44	2592.51
208	5/23/2011	21.58	2614.09
209	5/24/2011	22.36	2636.45
210	5/25/2011	21.32	2657.77

Appendix 16. Glycerol Treatment Reactor Biogas Production Data continued

Day	Date	Biogas Production Rate	Accumulated Biogas Volume
		l/day	l
211	5/26/2011	19.70	2677.47
212	5/27/2011	21.71	2699.18
213	5/28/2011	25.48	2724.66
214	5/29/2011	20.02	2744.68
215	5/30/2011	23.79	2768.47
224	6/8/2011	19.76	2788.23
225	6/9/2011	24.18	2812.41
226	6/10/2011	26.00	2838.41
227	6/11/2011	23.92	2862.33
228	6/12/2011	21.71	2884.04
229	6/13/2011	23.92	2907.96
230	6/14/2011	24.31	2932.27
231	6/15/2011	26.52	2958.79
232	6/16/2011	24.57	2983.36
233	6/17/2011	25.22	3008.58
234	6/18/2011	27.95	3036.53
235	6/19/2011	26.13	3062.66
236	6/20/2011	28.60	3091.26
237	6/21/2011	31.59	3122.85
238	6/22/2011	31.85	3154.70
244	6/28/2011	40.04	3194.74
245	6/29/2011	38.35	3233.09
246	6/30/2011	37.31	3270.40
271	7/25/2011	33.54	3303.94
273	7/27/2011	25.09	3329.03
274	7/28/2011	23.40	3352.43
275	7/29/2011	30.03	3382.46
276	7/30/2011	28.08	3410.54
277	7/31/2011	27.04	3437.58
278	8/1/2011	23.92	3461.50
279	8/2/2011	26.39	3487.89
280	8/3/2011	28.99	3516.88
281	8/4/2011	28.99	3545.87

Appendix 17. Two Point VFA Titration Data for figure 27

Day	Date	Control Reactor mg / l HAC	Glycerol Treatment Reactor mg / l HAC
118	2/22/2011	233	1,901
120	2/24/2011	175	1,828
121	2/25/2011		2,050
124	2/28/2011		1,845
125	3/1/2011	135	1,668
127	3/3/2011		1,320
128	3/4/2011		1,209
131	3/7/2011		1,039
132	3/8/2011		972
134	3/10/2011		977
135	3/11/2011		917
138	3/14/2011		780
139	3/15/2011		740
146	3/22/2011		1,156
149	3/25/2011	95	1,315
153	3/29/2011	236	1,171
155	3/31/2011	318	546
159	4/4/2011		373
167	4/12/2011	399	274
169	4/14/2011		361
174	4/19/2011		172
176	4/21/2011		245
181	4/26/2011		289
183	4/28/2011		361
188	5/3/2011	196	128
190	5/5/2011		826
191	5/6/2011		54
195	5/10/2011	0	0
197	5/12/2011		22
202	5/17/2011		500
204	5/19/2011		409
205	5/20/2011		480
210	5/25/2011		187
211	5/26/2011		74
216	5/31/2011	155	166

Appendix 17. Two Point VFA Titration Data for figure 27 continued

Day	Date	Control Reactor mg / l HAC	Glycerol Treatment Reactor mg / l HAC
217	6/1/2011	25	0
218	6/2/2011	141	115
223	6/7/2011		0
224	6/8/2011	137	28
229	6/13/2011	169	0
230	6/14/2011	982	97
237	6/21/2011	129	566
238	6/22/2011	171	632
239	6/23/2011	199	688
243	6/27/2011	252	690
244	6/28/2011	163	619
245	6/29/2011	71	598
246	6/30/2011	99	601
257	7/11/2011		243

VITA

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