FUNDAMENTALS OF OPERATION OF THE

INDUCED BED REACTOR (IBR) ANAEROBIC DIGESTER

by

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A dissertation submitted in partial fulfillment
of the requirements for the degree

of

DOCTOR OF PHILOSOPHY

in

Biological Engineering

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ABSTRACT

Fundamentals of Operation of the Induced Bed Reactor (IBR)

Anaerobic Digester

by

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Utah State University, 2010

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Department: Biological and Irrigation Engineering

The Induced Bed Reactor (IBR) was developed at Utah State University to apply high-rate anaerobic digestion techniques to high solids content substrates. This technology has been successfully implemented at full-scale multiple installations in the United States and Canada as a waste treatment and energy production technology, but the physical processes necessary to further optimize the system were not well understood.

Bench scale IBRs were operated as anaerobic digesters at 35°C, 45°C, and 55°C under three organic loading rates and three corresponding hydraulic retention times. Reactor performance was monitored at steady state for residence time distribution and substrate reduction.

The results show that the IBR behaves as a retained biomass reactor with fluid mixing that most closely approximates Completely Stirred Tank Reactor (CSTR) behavior when operated under the study conditions. A compartment real CSTR model, incorporating elements of dead zone and bypass flow, appears to be the most appropriate representation of the data. Mixing is likely due to a combination of energy inputs from thermal gradients induced by heat flux through the reactors and reactor and shear rates induced by gas evolution in the sludge bed.

(152 Pages)
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CHAPTER I

GENERAL INTRODUCTION

The purpose of anaerobic digestion in wastewater treatment is to reduce waste sludge volume and activity by fermenting the waste in the absence of oxygen. Anaerobic digestion relies on the symbiotic relationship between two general classes of anaerobic microorganisms to catabolize carbonaceous substrates to the relatively stable and innocuous end products of biosolids and biogas (Bryant et al., 1967). Biogas, consisting of primarily methane (CH₄) and carbon dioxide (CO₂), is generated as the primary respiration and oxidation end product of the microbial activity. Biosolids are the remaining solid products and include microbial cell mass, metabolic byproducts of cellular activity, and the indigestible fraction of the influent.

Anaerobes have long doubling times relative to aerobic bacteria. Historically, this has meant that anaerobic digestion can require large treatment volumes and long retention times (20-40 d) to enable populations to reproduce and develop sufficient concentrations to allow them to reduce substrate and meet treatment goals. These constraints can be avoided by utilizing high-rate treatment processes where populations of microbes (active solids) are maintained in the digester vessel, and substrate is passed through the active solids for treatment. Methods for accomplishing this include separating the active solids from the effluent and recycling them to the digester, by directly utilizing an attached growth process, or through the use of an induced bed or blanket of solids that self-select to remain in the reactor as the substrate passes through (Gerardi, 2003; Lettinga et al., 1980; McCarty, 1981; Tchobanoglous et al., 2003). These processes generally require a relatively low suspended solids concentration in the influent to avoid dilution of the active biomass and to avoid operational problems in the digester (plugging, sedimentation, short circuiting, etc).

This reliance on differentiation of solids and substrate poses a problem in treatment of some wastes, particularly those with high concentrations of suspended solids like those associated with food processing wastes and livestock manures. In the early 1990s, Dr. Conly Hansen at Utah State University proposed a digester configuration, the Induced Bed Reactor (IBR) (Hansen and Hansen, 2005), which was designed to address the operational limitations of applying high rate treatment techniques to high strength, high solids substrates.
The IBR has been successfully implemented with over 4,000 m³ of installed capacity in the US and Canada. No mathematical model has been developed, however, that would permit scaling or predictive evaluation of IBR performance. The lack of understanding of hydrodynamic behavior and treatment effectiveness limits deployment of the technology to known substrates and reactor sizes. Finally, while it is generally known that increased temperatures in anaerobic digestion result in increased reaction rates and corresponding reductions in required reactor volumes, there are no data on operation of the IBR at temperatures outside the range of 25-35°C.

RESEARCH OBJECTIVES

The overall objective of this research was to develop a fundamental understanding of hydrodynamic and kinetic behaviors of high rate anaerobic digestion processes in vertical upflow reactors at mid to moderate HRTs. Specific research objectives were as follows:

1. Develop hydrodynamic model for the Induced Bed Reactor
   a. Investigate the contacting patterns of bench scale digesters (60L) running clean water at steady state. Reactors were evaluated at three temperatures, 35°C, 45°C, and 55°C.
   b. Investigate the contacting patterns of three bench scale digesters engaged in active anaerobic digestion at steady state. Reactors were evaluated at three temperatures, 35°C, 45°C, and 55°C.

2. Investigate behavior of high rate suspended growth reactors evaluated in part 1:
   a. Three bench scale digesters engaged in active anaerobic digestion at steady state. Reactors were evaluated at three temperature levels (35°C, 45°C, and 55°C), and at three mass loading rates.

3. Recommend appropriate applications for the IBR in stabilization of high-strength wastes

LITERATURE REVIEW

DIGESTION AND WASTEWATER TREATMENT: PURPOSES

Wastewater is the liquid emission produced by a community or process when water has been utilized for some purpose, and has been contaminated to the point that it is no longer directly available for further use in that community or process. It may be contaminated with any number of organic or inorganic
constituents that exhibit as suspended, dissolved, or colloidal solids. Wastewater treatment is the process whereby that water is stabilized and cleaned for reuse or return to the environment. In biological treatment, microorganisms in the wastewater or receiving waters will act to remove organics and other constituents from solution via metabolism to new cellular material, or catabolism to energy and smaller molecules. In either case the carbon is separated from the wastewater as either gas or solid material, and the water is stabilized proportionally. This process of stabilization is commonly called digestion.

Depending on the environment in the water, digestion may be classified as aerobic, where sufficient $O_2$ is available in the environment to serve as the terminal electron acceptor and the substrate is completely oxidized to $CO_2$ and water, or anaerobic, where oxygen is not the terminal electron acceptor, and another terminal electron acceptor is used.

Equations 1 and 2 present the comparative energetics of oxidation using $O_2$ and $CO_2$ respectively as the terminal electron acceptors in the oxidation of acetic acid. In either case, approximately 40% of the energy will be made available to the cell to create more biomass or to sustain the cell, and 60% of the energy will be released to the environment as heat but in the aerobic process, the energy ($\Delta G_r$) released is significantly more than is made available in the anaerobic process.

\[
\frac{1}{8}CH_3COO^- + \frac{1}{4}O_2 \rightarrow \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + \frac{1}{4}H_2O \quad \Delta G_r = -105.82 \frac{Kj}{mol \ e^-} \quad (1)
\]

\[
\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O \rightarrow \frac{1}{8}CH_4 + \frac{1}{4}HCO_3^- \quad \Delta G_r = -3.55 \frac{Kj}{mol \ e^-} \quad (2)
\]

Both processes start with the same amount of chemical energy bound in the substrate molecule. In this context the principal difference between these processes is that an aerobic process will accumulate cell mass (biosolids or sludge) as the primary separable byproduct. In anaerobic respiration, significantly less energy is released and made available for cell growth. The anaerobic process will generate $CO_2$ and $CH_4$ as the primary products, and relatively less biomass. The balance of the energy from the substrate is bound in the methane that is a byproduct of anaerobic processes.

Aerobic processes convert carbonaceous substrates rapidly to cell mass and are regarded as stable and well understood, but they also require injection of large quantities of oxygen to maintain aerobic
conditions, leading to high power costs and infrastructure requirements. The rapid conversion of dissolved
and colloidal carbon to separable biomass creates a large volume of potentially regulated biosolids
requiring subsequent dewatering and disposal at an approved facility.

Anaerobic processes generate significantly less biosolids than aerobic processes, converting most of the
carbon instead to biogas. Biogas is typically 65% methane and 35% CO₂, with a heating value of
approximately 25 MJ/m³ at that ratio. The energy bound in methane biogas can be recovered and used for
process or other heating requirements (Gerardi, 2003; Speece, 1996; Tchobanoglous et al., 2003).
Depending on the process design, the relatively slow doubling times can lead to long hydraulic retention
times for anaerobic treatment with corresponding increases in infrastructure costs. High rate systems are
sensitive to loading rates and care must be taken to maintain process parameters in a range acceptable to
the microbial communities (Gerardi, 2003; Tchobanoglous et al., 2003). Finally, anaerobic digesters are
sometimes less well understood by operators, leading to a lack of confidence in the technology.

ANAEROBIC DIGESTION REACTOR DESIGN:
HISTORICAL APPROACH

Scientific investigation of anaerobic digestion for waste volume reduction can be dated to 1881, when
M. Mouras published a description of his “Automatic Scavenger,” a sealed container in which organic
wastes were liquefied by anaerobic decomposition (McCarty, 1981, 2001). In 1894, the Massachusetts
State Board of Health issued a report that recognized sludge destruction in septic tanks was dependent on
slow microbial action, and thus recommended cleaning only when necessary to maintain healthy bacterial
populations (McCarty, 1981). The first septic tank was patented by Donald Cameron in Exeter, England in
1895, and implemented for the pretreatment of wastewater in Exeter and in the US at Urbana and
Champaign Illinois. Mr. Cameron’s observations led to the capture and reuse of the methane produced in
the tanks for heat and lighting at the Exeter facility.

The septic tanks produced an offensive sludge; in 1899, Harry Clark proposed capturing the sludge and
fermenting it separately. Building on work by William Travis, Karl Imhoff constructed the first successful
two-stage system in which a two-compartment vessel provided for solids separation in the upper chamber,
and retention in a lower chamber. This permitted long solids retention times in a relatively small volume,
and the concentrated sludge was fermented over a period of weeks or months until it was stabilized to the point that it could be discharged without nuisance.

By the 1920s, operators and researchers had begun to realize the advantages of multi-stage systems, utilizing clarifiers to separate the liquid and solid fractions of wastewaters, and drawing off the settled sludges for separate digestion. In 1927, a German treatment facility installed the first heated sludge digestion tank. Control of process temperature improved performance significantly over Imhoff Tanks, and separate digestion became increasingly the technique of choice.

In 1927, it was shown that while the quantity of gas produced by a given amount of sludge was constant, the rate of digestion was directly proportional to temperature. In the early 1930s, two temperature optima were determined for anaerobic digestion, one in the mesophilic range, and one in the thermophilic range. Thus by the 1930s, the following basic conditions had been established for optimization of anaerobic digestion of sludges:

1) Concentrate sludge
2) Control temperature
3) Maintain an anaerobic environment
4) Potential gas production volume is absolute for a given quantity of sludge
5) Gas production rate is directly proportional to process temperature

The next major breakthrough in the development of digestion technology took place with when G. J. Stander recognized the benefits of maintaining high microbial population concentrations in digester vessels. By separating the active solids from the effluent stream and maintaining them in the digestion vessel, he was able to substantially reduce detention times for wastewaters (Stander and Snyder, 1950). Independently, researchers working with dilute industrial wastes achieved similar results by using a separate vessel for sedimentation of digester effluents, then recycling the solids back to the digester. By doing so, they were able to reduce detention times by a factor of twenty (Schroepfer et al., 1955).

Another significant advance in the 1950s was the advent of mechanical mixing. Without mixing, digester contents form three phases, a solid sludge, a liquid supernatant, and an aerated floating scum layer. Researchers showed that by mechanically mixing the digester contents, they could redistribute the reactor contents, removing the scum layer and bringing the active solids concentrated in the sludge in
contact with the dissolved substrate in the supernatant. Not only did this improve reactor efficiency by eliminating the scum layer and making the full reactor volume available for treatment, it improved the conversion performance significantly as well (Morgan, 1954).

From these developments, designers were able to begin to optimize the mechanical design of digesters. By this time, they knew that they wanted to maintain a constant temperature and an anaerobic environment. They also knew that it was more effective to recycle concentrated active sludge than to grow it up new for each batch of substrate, and that a high concentration of active sludge coupled with good distribution of substrate (mixing) would result in improved treatment conditions. With the work of Bryant, et al (1967), they also had a reasonable basis for understanding the microbiology of anaerobic digestion, and particularly the distribution and interaction of the three phases of the process and the environmental requirements of the microbial communities doing the work.

Researchers began to develop reactors designed to address these parameters. Among the first was Stander’s “clarigester” (1966), an upflow reactor where substrate was introduced into an unmixed chamber in the bottom of the reactor, percolated through the active solids up into a second chamber where any suspended solids would settle and return to the bottom of the reactor, and supernatant from the upper chamber would flow out of the system. Other developments followed including attached growth anaerobic filters (similar to aerobic trickling filters), anaerobic attached film expanded bed reactors, anaerobic baffled reactors, and upflow anaerobic sludge blanket reactors (Lettinga et al., 1980; Switzenbaum and Jewell, 1980; Taylor, 1972; Young and McCarty, 1969). These approaches, in conjunction with traditional completely stirred tanks reactors (CSTRs) with sludge recycle, gave designers a range of acceptable options to work with in treatment of high-strength wastes, and validated the work done previously to develop an understanding of the factors that were important to system function.

**Anaerobic Reactor Design: Current Practice**

There are two approaches in the chemical engineering literature to reactor design; the petrotech approach, and Chemical Reaction Engineering (CRE) (Levenspiel, 1999a). The first comprehensive attempt at systematizing the relatively new discipline of chemical engineering came in 1947 with *Chemical Process Principles: Part III, Kinetics and Catalysis* (Watson and Hougen). This text
essentially codified the petrotech approach which relies on deterministic analysis of each reaction anticipated in a system, and the subsequent determination of the rate limiting reaction. While it is a rigorous way of addressing the problem, it requires identification of all the variables in a system, and design of experiments adequate to quantify each of them. For complex biological systems with varying substrate characteristics and evolving microbial communities, this is simply not practical.

A useful model must be general enough to be broadly applicable for the anticipated range of reaction conditions. A reactor model has to account for both kinetics and flow patterns, and can only be as reliable as its least well defined component.

The principles of chemical reaction engineering permit development of simple kinetic rate expressions that do not attempt to define specific mechanisms, but to look at the system as a whole. Coupling this kinetic approach with quality work to determine contacting and flow patterns in reactors gives designers a powerful tool for developing general reactor design equations. That’s not to say that there is less value in doing the work that enables determination of specific rate expressions and mechanisms; on the contrary, such work is what reveals the basic building blocks to permit better understanding of process fundamentals. For parameter development to permit scaling and design, however, the CRE approach is most appropriate.

**Chemical Reaction Engineering**

Chemical reaction engineering (CRE) combines the study of chemical kinetics with the reactors where they take place (Fogler, 2006). By developing an appropriate understanding of flow pattern and kinetics, an equation that explains reactor performance can be developed (Levenspiel, 1993). The performance equation (Figure 1-1, Eq. 3) can then be used as a basis to scale the reactor or predict performance under changed conditions.

\[
\text{Output} = f(\text{input, kinetics, contacting})
\] (3)
Figure 1-1: Chemical Reaction Engineering model

Equation 3 is essentially a material mass balance; it provides the fundamental basis for reactor design. Analysis of any reactor can be approached by developing a mass balance to relate input to output, and understanding the interplay between contacting patterns and kinetics as they impact the conversion of mass in the system and thus the effluent composition. This can be applied to analyzing the influent in sufficient detail to understand its impact on the process, then constructing the fundamental mass balances required to define the contacting pattern in the reactor and thus the efficacy of distribution of both the active solids catalyst and the substrate, and developing a kinetic model appropriate for the chemical reactions taking place.

Influent Characterization

For any anaerobic digestion process, there are a common set of wastewater characteristics that have been demonstrated over time to be important as outlined in Table 1-1.

Evaluation of influent will include analysis of each of these components to determine their impact on the desired outcome, and will also impact the approach taken to digestion. If, for example, there is a high fraction of undissolved organic material, de-coupling of SRT and HRT becomes more difficult as the undissolved solids may have a tendency to accumulate in the reactor; without preliminary treatment to remove the solids, a simple CSTR batch system might be appropriate. For a waste with similar strength but a completely dissolved substrate, a high rate digester with a substantially lower HRT might be acceptable. Assuming a 20-day HRT for the first instance and a 0.5 day HRT for the second, the completely dissolved substrate achieves the desired result with only 2.5% of the volume. By the same
token, if a high rate suspended growth system is selected for the first substrate, it will accumulate solids and cease to function as designed.

Table 1-1: Wastewater characteristics for anaerobic treatment processes (adapted from Tchobanoglous et al., 2003)

<table>
<thead>
<tr>
<th>Flow/Loading Variations</th>
<th>Microbial complexity requires consistent flow and loading conditions to avoid process upsets. Provide flow equalization or additional capacity to attenuate potential shock loadings. Hydraulic Retention Time (HRT) is a critical process parameter.</th>
</tr>
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<td>Organic concentration and temp.</td>
<td>Optimal temperatures are between 25-35°C and 50-65°C for anaerobic digestion. Most digesters operate in the mesophilic range, although reaction rates as tracked by methane production double for every 10°C increase in temperature. Consider aerobic treatment for COD loading below 1500 mg/L. Organic Loading Rate (OLR) is a critical process parameter.</td>
</tr>
<tr>
<td>Fraction of Non-Dissolved Organic Material</td>
<td>Wastewaters with high solids fractions better suited for suspended growth processes than attached growth to prevent plugging. If greater conversion of particulate organic matter is required, long SRT values may be needed to accommodate hydrolysis as rate limiting step.</td>
</tr>
<tr>
<td>Wastewater Alkalinity</td>
<td>To maintain optimal pH for methane production, high levels of carbonate alkalinity are required in digesters as CO₂ production constantly removes alkalinity from the system. If chemical addition is required for pH control, this can impact economics.</td>
</tr>
<tr>
<td>Nutrients</td>
<td>Although anaerobic digestion does not produce large quantities of biomass, some industrial feedstocks may lack the nutrients required to support growth. Anaerobic digestion will not remove significant quantities of nutrients as biomass.</td>
</tr>
<tr>
<td>Toxic Compounds</td>
<td>Appropriate toxicity studies should be conducted on the substrate to ensure that chronic toxicity does not exist. See Parkin and Owen (1986) for a comprehensive discussion. Anaerobic cultures have the capacity to acclimate to some potentially inhibitory substances (Speece, 1996).</td>
</tr>
<tr>
<td>Solids Retention Time (SRT)</td>
<td>SRT is a fundamental design and operating parameter. In general, SRTs of &gt;20 days are required for effective treatment performance at 30°C.</td>
</tr>
<tr>
<td>Expected Methane Gas Production</td>
<td>The amount of methane produced under anaerobic conditions is 0.35 L CH₄/g COD @ 0°C and 1 atm. If methane is to be recovered, this parameter is important in economic evaluation of the system. Methane can also be used to monitor process performance.</td>
</tr>
<tr>
<td>Treatment Efficiency Required</td>
<td>Anaerobic treatment processes are capable of high COD conversion efficiencies, but discharges may exceed effluent limitations requiring additional downstream processing.</td>
</tr>
<tr>
<td>Sulfide Production and Toxicity</td>
<td>Excessive oxidized sulfur in the influent can result in high concentrations of hydrogen sulfide gas and hydrogen sulfide (HS⁻) and sulfide (S²⁻) ions in solution. Sulfide can be toxic in methanogenic systems. H₂S is corrosive and malodorous.</td>
</tr>
<tr>
<td>Ammonia Toxicity</td>
<td>Ammonium can be toxic to methanogenic systems, but given steady state operation and time to acclimatize, ammonia toxicity has been shown to be something cultures can adapt to. Consideration should be given to potential for ammonia formation and shock loadings (Angelidakis et al., 1993).</td>
</tr>
<tr>
<td>Liquid-Solids Separation</td>
<td>Effective separation correlates directly to higher solids retention time (SRT). Given the very low biomass production rate and the critical function of active solids in catabolizing substrate, the higher the SRT, the better the treatment efficiency of the reactor.</td>
</tr>
</tbody>
</table>
Contacting and Flow Patterns (Mixing)

Reactors can be classified by their hydraulic properties. There are two ideal reactor types for steady state conditions; completely stirred tank reactors (CSTR) and plug flow reactors (PF). In an ideal CSTR reactor, the concentration of all constituents is equal at all points in the reactor; as soon as something is introduced at the inlet, it is immediately dispersed throughout the total reactor volume. The ideal PF reactor is the opposite; influent enters, crosses the reactor volume, and exit as a discrete packet with no mixing. These reactors are mathematically predictable, and therefore easy to analyze, but one of them will also generally provide the best outcome for a desired reaction.

In reality, no reactor will meet these ideals perfectly, but in a well designed reactor, some particles will be retained longer than the average and some will pass out of the reactor more quickly but statistically the average particle will remain in the reactor for the amount of time required to complete the design reaction.

The fundamentals of reactor characterization found their origins in MacMullin and Weber’s 1935 approach to stirred tank reactors in series, but in his classical 1953 paper, Danckwerts laid the groundwork and defined the nomenclature for the approach that has become the standard for reactor analysis (Danckwerts, 1953; MacMullin and Weber, 1935).

Reactor characterization is addressed using statistical methods to determine the average amount of time a particle spends in a reactor and the shape of the distribution. The shape of the distribution can also provide information on the volumetric efficiency of the reactor by identifying bypassing, channeling, and dead space. Other authors have made refinements to Danckwerts’ techniques (Coker, 2001; Levenspiel, 1993, 1999b; Tchobanoglous et al., 2003; Wen and Fan, 1975) but the fundamental approach as outlined below has stood the test of time.

Danckwerts’ method is founded on determining the residence time distribution (RTD) in a reactor. An accurate picture of RTD allows 1) determination of contacting and flow patterns in the reactor, which in turn permits accurate linkage of those patterns to kinetics, and 2) verification of contacting and flow patterns when reactors are scaled. Consequences of failure to adequately represent mixing include wasted reactor volume with associated costs and in the case of a wastewater facility, inability to meet treatment requirements.
Experimental Determination of Residence Time Distribution

Residence time distribution (RTD) is a statistical characterization of the time that particles spend in a reactor, and the most important factor in reactor characterization (Wen and Fan, 1975). Form in reactor design is much less important than function; similar reactor behaviors can be derived from very different geometries if they have the same RTDs since similar contacting patterns will yield similar performance equations. This is particularly significant in the scalability of systems.

RTD is determined by conducting tracer studies where a tracer of known concentration is injected into the inlet of a reactor, and the concentration in the reactor fluid is measured at points of interest, usually the outlet.

Tracer Selection

Selection of a tracer that will provide the data required for the analysis is critical. The following characteristics for tracer selection are adapted from Tchobanoglous (2003) and Denbeigh and Turner (1965). An appropriate tracer:

1. Will have no impact on flow (e.g. same density as fluid when in solution)
2. Is conservative to enable a mass balance
   a. Should not react with or sorb to reactor surfaces or contents
3. Is injectable over a relatively short time period
4. Is convenient to analyze
5. Has low molecular diffusivity

Inlet and outlet conditions are also critical (Levenspiel, 1993); the tracer must be injected appropriately so as to ensure that results are not skewed by non-uniform application to the reactor contents. The tracer injection should match as closely as possible the normal steady state influent stream. Measurement of the effluent can be accomplished in one of three methods: grab, “mixing cup,” or in-line samples taken at recorded intervals. Grab samples capture a discrete data point of concentration, and depending on the reactor mixing characteristics may provide considerable scatter in the data. Without a sufficiently dense sampling program, this may provide data that is difficult to interpret accurately, and if the sampling interval is sufficiently large, may miss important inflection points altogether. Mixing cup measurements
are generally preferred as they capture the full volume of the reactor effluent over the sampling interval and provide an average value for the effluent concentration over the time of the measurement. The principle difficulty in working with mixing cup measurements is in capturing, storing, and mixing the reactor effluent through the sampling interval.

A third option is in-line monitoring as is possible with radioactive tracers (Borroto et al., 2003; Samson and Guiot, 1985; Wen and Fan, 1975) or in situ fluorescence based analysis (Lou et al., 2006). The advantages are that there is no limit on the number of data points that can be collected beyond the sampling frequency of the data acquisition system, and the behavior of the system can be monitored at multiple points axially and radially within the process to get a more accurate picture than a simple outlet concentration will provide. This in turn provides direct evidence regarding earliness or lateness of mixing, dead zones, and other reactor characteristics which otherwise have to be inferred indirectly from RTD curve analysis. On the other hand, radiotracers require specialized controls and carry licensing requirements and both radiotracer and in situ fluorescence require specialized monitoring equipment that may not be justifiable for many analyses.

Tracers selected for this work were rhodamine WT and lithium chloride. Rhodamine WT is a fluorescent dye that has been used successfully in anaerobic digestion studies (Lou et al., 2006; Tchobanoglous et al., 2003). It is relatively inexpensive when purchased in bulk, it is simple to analyze as a fluorophore, and it meets the other requirements for a good tracer. There is some concern in the literature regarding an isomer of rhodamine WT that may sorb to surfaces (Vasudevan et al., 2001). The principal advantage to a fluorescent tracer is in ease of analysis; by using a plate reader fluorimeter, up to 96 samples can be evaluated consistently and rapidly at minimal cost.

One potential difficulty in working with a fluorescent tracer is that use of fluorimetry yields an indirect measurement that must be correlated to concentration. In order to close the mass balance conformational studies utilizing direct measurements were required. Three confirmatory tracer studies were performed using lithium chloride as the tracer and monitoring the effluent for Li⁺ (Leighton and Forster, 1996; Olivet et al., 2005). Lithium can be inhibitory, and care must be taken in the experimental design to ensure that excessive amounts are not injected, particularly if a pulse function tracer injection is used (Anderson et al., 1991).
**Tracer Injection**

The first physical step in a tracer study is the injection of the tracer. There are several approaches available to the researcher including pulse, step, periodic, and random inputs. For most purposes, the pulse and step input functions are adequate.

Pulse input involves a single injection of tracer over a short time period relative to the HRT of the system; the objective is to have an instantaneous input. By monitoring the outlet concentration of a pulse tracer input and plotting the data vs. time, a curve will be generated that represents the residence time distribution of the fluid in the reactor. This is referred to as the C curve. Figure 1-2 provides a graphical summary of the input conditions and potential response curves for slug input of tracer in both CSTR (a-1) and PF (b-1) systems.

Step input involves a sudden shift in concentration of the tracer at the inlet to the reactor. Step input studies are more difficult to set up in general because of the requirement to maintain a constant concentration in the feed stream, and particularly in large scale systems, the dependence on other processes for steady state operation and the relatively large quantities of tracer required. The chief advantage of the step function is that when the reactor contents have completely turned over (i.e. been replaced by the tracer feed) it is immediately evident in the effluent concentration measurements; there is no danger of underestimating the peak as might happen with a step function. It is also easier to close the mass balance on a step input function as the test is run until \(C_{out} = C_{in}\). Figure 1-3 provides a graphical summary of the input conditions and potential response curves for slug input of tracer in both CSTR (a-2) and PF (b-2) systems.

**Tracer Analysis**

Recovered samples are analyzed to determine tracer concentration. Concentration (C) is plotted as a function of time; this is known as the C-t curve (Figures 2, 3). The curve is typically then normalized to provide a unitless representation of the flow pattern. Time is normalized to the mass balance calculated HRT, and concentration is normalized to \(C_{in}\), the influent concentration. The normalized time value is plotted as \(\theta\), while the normalized concentration is presented as either \(E\) for a pulse input or \(F\) for a step input. \(E\) represents the exit age distribution function, while \(F\) represents the cumulative exit age.
distribution function (Figure 1-4). The F-curve is the integral of the E-curve; thus, regardless of which experiment is chosen, both curves can be derived from the output. It should be noted however that since the E-curve is the slope of the F-curve, it magnifies experimental error, and thus the F-curve derived directly from a step input experiment will give more accurate results than the E-curve developed from a pulse input function.

Figure 1-2: Slug or pulse input of a tracer and anticipated C-T curves (adapted from Tchobanoglous et al., 2003)

Figure 1-3: Step input of a tracer and anticipated C-T response curves (adapted from Tchobanoglous et al., 2003)
Figure 1-4: Comparison of E- and F-Curves for a non-ideal plug flow reactor (adapted from Tchobanoglous et al., 2003).

The usefulness of the curves is in establishing the behavior of the reactor; from them, one can quickly see how closely the reactor behavior approached the ideals of CSTR and PF. The plots will also indicate volumetric inefficiencies like dead space, short circuiting, and channeling.

Given that the E-curve represents the exit age distribution, the first moment of the curve will represent $\tau$, the experimentally derived hydraulic retention time. The variance, $\sigma^2$, indicates the spread of the distribution. These values can be directly applied to compare the distribution to idealized responses and determine the appropriate model for the reactor in question. Another number that can be useful in defining reactor performance is the dispersion factor, $d$. $d$ is the inverse of the Peclet number, $P_e$, and can be used to evaluate degree of mixing in a reactor (Table 1-2). Derivation and interpretation of the curves is addressed in detail in the literature (Coker, 2001; Danckwerts, 1953; Fogler, 2006; Levenspiel, 1993, 1999b; Wen and Fan, 1975). In wastewater, several parameters derived from the E-curve can be applied to further classify the reactor performance and to compare reactors directly (Table 1-3).
Table 1-2: Degree of dispersion (adapted from Tchobanoglous et al., 2003)

<table>
<thead>
<tr>
<th>d</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>d=0</td>
<td>No dispersion (plug flow)</td>
</tr>
<tr>
<td>d&lt;0.05</td>
<td>Low dispersion</td>
</tr>
<tr>
<td>0.05≤d≤0.25</td>
<td>Moderate dispersion</td>
</tr>
<tr>
<td>d≥0.25</td>
<td>High dispersion</td>
</tr>
<tr>
<td>d→∞</td>
<td>CSTR</td>
</tr>
</tbody>
</table>

Table 1-3: Descriptive terms for use in classifying wastewater reactors from RTD E-Curve data (adapted from Danckwerts, 1953, Tchobanoglous et al., 2003, Levenspiel 1993)

<table>
<thead>
<tr>
<th>Term</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>τ</td>
<td>Theoretical hydraulic residence time (HRT) (volume/volumetric flow rate)</td>
</tr>
<tr>
<td>t₁</td>
<td>Time at which tracer first appears</td>
</tr>
<tr>
<td>tₚ</td>
<td>Time at which peak tracer concentration is observed</td>
</tr>
<tr>
<td>t₄₋ₚₚ₋ₚₙ</td>
<td>Time to reach centroid of RTD (observed HRT)</td>
</tr>
<tr>
<td>t₉₀/t₅₀,t₉₀</td>
<td>Time for 10, 50, and 90% of the tracer to pass through the reactor</td>
</tr>
<tr>
<td>1/MDI</td>
<td>Morrill Dispersion Index (MDI)</td>
</tr>
<tr>
<td>τ/τ</td>
<td>Volumetric efficiency as defined by Morrill</td>
</tr>
<tr>
<td>t₉₀/τ</td>
<td>Index of short circuiting. In an ideal PF, the ratio is 1 and approaches zero for a CSTR</td>
</tr>
<tr>
<td>t₉₀/τ</td>
<td>Index of modal retention time. Ratio will approach 1 in an ideal PF and 0 in a CSTR. For values of the ratio greater than or less than 1.0, flow distribution in the reactor is not uniform.</td>
</tr>
<tr>
<td>t₉₀₀/τ</td>
<td>Index of average retention time. A value of 1 indicates full use of the volume.</td>
</tr>
<tr>
<td>σ²</td>
<td>Variance of the distribution; indicates the spread of the distribution</td>
</tr>
</tbody>
</table>

MICROBIOLOGY OF ANAEROBIC DIGESTION

In 1967, Bryant (Bryant et al., 1967) showed that anaerobic digestion can be summarized as a three-stage process requiring the syntrophic activity of three distinct groups of organisms (Figure 1-5, Figure 1-6) in an oxygen-free environment. The stages are mutually dependent and sequential. First, complex and colloidal organics are hydrolyzed into either intermediates or directly to methane precursors. Second, the intermediates are fermented to methane precursors. Finally, via methanogenesis, methane and carbon dioxide are produced.

The anaerobic digestion process works in a defined sequence, where the products of one group become the feed substrate for the next (Gerardi, 2003; Zeikus, 1981). Anaerobic digestion can only proceed smoothly if these stages are maintained at stoichiometrically balanced rates. If any stage is inhibited or in excess, it will impact the others. For example, if the hydrolysis stage is inhibited, it will result in a shortage of substrates for acidogenesis and production of volatile fatty acids (VFAs). Without VFAs, methane production will decrease. By the same token, if VFAs accumulate, they will consume alkalinity, depressing pH, and the methanogenic activity will decrease. The rate limiting step in anaerobic digestion
to methane is generally production of acetate, although depending on influent characteristics, hydrolysis of complex molecules may be limiting (Gerardi, 2003).

![Diagram of anaerobic process schematic of hydrolysis, fermentation, and methanogenesis](image1)

**Figure 1-5:** Anaerobic process schematic of hydrolysis, fermentation, and methanogenesis (adapted from Tchobanoglous et al., 2003)

![Diagram of carbon and hydrogen flow in anaerobic digestion process](image2)

**Figure 1-6:** Carbon and hydrogen flow in anaerobic digestion process (adapted from Tchobanoglous, Burton et al 2003).

In the undefined cultures found in anaerobic digestion, a wide variety of organisms can fill these niches including many species of facultative anaerobes, strict anaerobes, and archaea. The precise species
are not as important as the gene expression which permits utilization of the available substrate and conversion to facilitate each stage in the process. This is convenient in that given sufficient time and influent consistency, cultures will adapt to conditions and select for the most effective communities.

Understanding the interactions between the populations in an anaerobic digestion system and their biochemical and environmental requirements allows determination of optimal system design and operation. Given that digestion requires the maintenance of a balance between three distinct processes, and the maintenance of diverse microbial communities with differing requirements, it follows that the design and operation should be geared toward providing an environment that satisfies the needs common to all three and vital to each.

**Hydrolysis**

Not all substrates require hydrolysis and fermentation; acetate, H₂ and CO₂, methanol, formate, and some other compounds may be found in substrates and can be directly catabolized to methane. Most wastewaters are by definition complex, however, and contain a mixture of insoluble carbon compounds including lipids, polysaccharides, proteins, and nucleic acids that must first be split into simpler molecules to permit them to be directly utilized inside the cells. In this process, hydrolytic bacteria release extracellular enzymes which add water to the chemical bonds between complex molecules, breaking the bonds and allowing the constituent molecules to go into solution. Once hydrolyzed, the simple molecules in solution can be transported across cell membranes for utilization.

The hydrolysis step itself is a conversion of carbon forms, and does not constitute extraction of carbon from the waste and therefore to the stabilization of the wastewater. It simply makes the carbon available to the cells that will actually utilize the material.

**Acidogenesis**

In the second stage of anaerobic digestion, the hydrolyzed organics are progressively fermented from relatively long chain acids and sugars to small organic acids like acetic, propionic, and butyric acids, and to carbon dioxide, hydrogen gas, alcohols, and organic compounds containing nitrogen and sulfur (Gerardi, 2003; Parkin and Owen, 1986). Table 1-4 shows the major components produced through fermentation processes in anaerobic digesters and their usefulness as substrates for methane formation.
Table 1-4: Major products of acidogenesis and their suitability as substrate for methanogenesis (adapted from Gerardi, 2003)

<table>
<thead>
<tr>
<th>Product</th>
<th>Formula</th>
<th>Substrate for Methanogenesis?</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetate</td>
<td>CH₃COOH</td>
<td>Direct</td>
</tr>
<tr>
<td>Butanol</td>
<td>CH₃(CH₂)₂CH₂OH</td>
<td>No</td>
</tr>
<tr>
<td>Butyrate</td>
<td>CH₃(CH₂)₂CH₂COOH</td>
<td>Indirect</td>
</tr>
<tr>
<td>Caproic Acid</td>
<td>CH₃(CH₂)₃COOH</td>
<td>No</td>
</tr>
<tr>
<td>Formate</td>
<td>HCOOH</td>
<td>Direct</td>
</tr>
<tr>
<td>Ethanol</td>
<td>CH₃CH₂OH</td>
<td>Indirect</td>
</tr>
<tr>
<td>Lactate</td>
<td>CH₃CHOHCOOH</td>
<td>No</td>
</tr>
<tr>
<td>Methanol</td>
<td>CH₃OH</td>
<td>Direct</td>
</tr>
<tr>
<td>Propanol</td>
<td>CH₃CH₂CH₂OH</td>
<td>No</td>
</tr>
<tr>
<td>Propionate</td>
<td>CH₃CH₂COOH</td>
<td>Indirect</td>
</tr>
<tr>
<td>Succinate</td>
<td>HOOCCH₂CH₂COOH</td>
<td>No</td>
</tr>
<tr>
<td>Methalmine</td>
<td>CH₃NH₂</td>
<td>Direct</td>
</tr>
<tr>
<td>H₂, CO₂</td>
<td>--</td>
<td>Direct</td>
</tr>
</tbody>
</table>

Acetic acid is the most important of these compounds in terms of its utilization by methanogens to generate methane, accounting for 72% of the methane produced (Figure 1-6). Acetic acid may be formed by acidogenic bacteria, hydrogen consuming acetogens, and hydrogen producing acetogens. The acetogenic bacteria can convert the more complex compounds like butyrate, propionate, and alcohols to acetate which can then be used by the methanogens. Some of the more complex carbon compounds require further fermentation to carbon dioxide and hydrogen to generate methane precursors.

The concentration of hydrogen in an anaerobic digester is an important factor in process stability; as long as the partial pressure of hydrogen is less than 10⁻¹ atmospheres, acetic acid formation is favored and methane is the major product. Above this level, production of propionic acid, butyric acid, and ethanol are favored. This highlights the importance of the syntropic relationship between the acid forming and methane producing communities; when the populations are balanced such that the methanogens can remove H₂ effectively as methane substrate and maintain a low partial pressure of hydrogen, the acid formers can produce primarily acetate, which then helps remove the hydrogen. And so it goes.

**Methanogenesis**

Methanogenesis is the stage in which waste is actually stabilized as carbon is removed from the system as CO₂ and CH₄. Methane is essentially insoluble in water, and self separates as a gas. CO₂ either escapes as gas or is incorporated as bicarbonate alkalinity.
Methanogenesis is accomplished by strictly anaerobic archaea. It is produced via two major pathways; first, methylotrophic methanogens produce methane from methyl-group containing substrates like methanol, methalamine, and acetate (Eq. 4, 5, 6) by splitting off the methyl group. As noted previously, acetate is by far the most significant pathway, responsible for 72% of methane formation.

\[ 2\text{CH}_3\text{OH} + 3\text{H}_2 \rightarrow 3\text{CH}_4 + 3\text{H}_2\text{O} \quad \text{(Methanol)} \quad (4) \]

\[ 4(\text{CH}_3)_3\text{N} + 6\text{H}_2\text{O} \rightarrow 9\text{CH}_4 + 3\text{CO}_2 + 4\text{NH}_3 \quad \text{(Methylamine)} \quad (5) \]

\[ \text{CH}_3\text{COOH} \rightarrow \text{CH}_4 + \text{CO}_2 \quad \text{(Acetate)} \quad (6) \]

Chemolithotrophic methanogens are able to form methane primarily from carbon dioxide and hydrogen (Eq. 7), accounting for the balance of the methane produced (28%), although formate can also be split to yield methane and water (Eq. 8).

\[ \text{CO}_2 + 4\text{H}_2 \rightarrow \text{CH}_4 + 2\text{H}_2\text{O} \quad \text{(Carbon Dioxide and Hydrogen)} \quad (7) \]

\[ 2\text{HCOOH} \rightarrow \text{CH}_4 + \text{H}_2\text{O} \quad \text{(Formate)} \quad (8) \]

The methanogenic archaea grow much more slowly than the acid forming bacteria that produce their substrate. They are also more sensitive to environmental conditions. Methane production is therefore generally the rate limiting step in anaerobic digestion systems (Parkin and Owen, 1986).

**MODELING OF ANAEROBIC DIGESTION**

"Everything should be made as simple as possible... but no simpler."

--Albert Einstein

In order to develop a rational model for a unit process, the dynamic interactions of substrate utilization and microbial growth must be understood. Over the past 100 years, significant advances have been made in the understanding of anaerobic digestion processes, first in terms of developing reactors that supported
natural processes, then by developing an understanding of the fundamental microbiological underpinnings of those processes (McCarty, 1981, 2001).

There are probably as many approaches to modeling of anaerobic digestion processes as there are digester types; the process itself is delightfully complex in its microbial underpinnings. This complexity lends itself to attempts by researchers to explore the extremes of modeling possibilities by either making simplifying assumptions or attempting to comprehensively define each interaction. On the simplification side, researchers have applied Monod, Mikales-Menton, and first order kinetics to develop lumped parameter steady state models with varying degrees of success (Chen et al., 1980; Garcia-Ochoa et al., 1999; Hashimoto, 1982; Husain, 1998; Karim et al., 2007; Rodriguez Andara and Lomas Esteban, 1999; Varel et al., 1980). Others have pursued complete process definition with complex general dynamic models (Angelidaki et al., 1993, 1999; Batstone et al., 2002).

Both approaches have merit, depending on their intended application. The dynamic models make a good platform for simulation of unsteady (startup) and cyclic processes. Steady state models are simple, easily adapted, and provide sufficient resolution to represent most processes. Both types also require a fundamental understanding of the process to be modeled as outlined below so that the user can be aware of the advantages and limitations provided by each, and select the method most appropriate for the task at hand.

For the purposes of this research, the approach to the kinetic model must be suitable for use in a steady state system and be flexible enough to address multiple temperature ranges, multiple substrates (viz. manure and artificial manure), and utilize COD and biogas as the principle measures of performance.

**Dynamic Distributed Parameter Models**

Distributed parameter models attempt to mathematically express the interactions and interdependencies between the microbial communities involved in anaerobic digestion by developing and integrating the mass balances for the distinct substrates and bacterial cultures involved and include consideration of inhibitory effects. Two models that have been developed along these lines are the model developed by the Danish Energy Research Program (DERP) for use with manures codigested with other substrates (Angelidaki et al., 1993, 1999), and the International Water Association Anaerobic Digestion Model 1
(ADM1) for use with complex wastewaters (Batstone et al., 2002). A third model was proposed by DT Hill at Auburn University for the dynamic modeling of animal wastes (1982).

The DERP model incorporates two enzymatic processes and eight distinct bacterial groups. It also incorporates six types of inhibition. The substrate is defined by dead cell mass, carbohydrates, proteins, lipids, and their degradation products. Inorganic constituents include ammonia, phosphate, carbon, hydrogen sulfide, anions, and cations. Carbohydrates are subdivided into soluble, insoluble, and inert fractions. Kinetic expressions rely on both first order (hydrolysis, cell decay) and Monod kinetics (primary growth and ammonia substrate dependency).

The model was validated against CSTR digesters fed manure and various codigestive substrates, and shown to yield results similar to observed data from full scale systems. The authors were comfortable in recommending it as a tool for prediction of process performance for digestion of complex wastes, but recommended further studies be done.

The ADM1 model was co-developed with the principal author of the DERP model, and incorporates some of its methodology but is considerably more complex. It incorporates 25 processes with their associated rate expressions and 19 components. The model is intended as a more general approach than the DERP model; studies utilizing ADM1 addressed its use for two stage mesophilic/thermophilic anaerobic digestion of municipal waste (Blumensaat and Keller, 2005), and a UASB reactor (Tartakovsky et al., 2007). Tartakovsky modified the model to incorporate lateral dispersion. Blumensaat noted that the nitrogen mass balance would not close without adjusting the model and that while the carbon mass balance would not close, the errors were acceptable. However, all researchers recommended the model for evaluation of anaerobic systems.

Hill’s model, like the DERP model, focuses on manure digestion and provides a simplifying approach to substrate availability by normalizing the substrate to a dimensionless biodegradability coefficient. Of the three dynamic models, Hill is by far the simplest. He reduces the interactions to an acidogenesis step and a methanogenesis step where the acidogens utilize the biodegradable portion of the waste as substrate while the methanogens utilize the VFAs produced by the acidogens. Even with these simplifications, it is still computationally intensive. In addition, its use is restricted to the substrates defined by the author.
Husain provided a steady state solution to Hill’s model for any substrate by redefining the inputs (Husain, 1998) as outlined in the Steady State section below.

Given their inherent complexity, researchers and designers should approach the use of dynamic models with caution and validate the results with experimental data. The greatest advantage to them is that they can be used to dynamically simulate non-steady state behavior including changes in feed composition or strength, inhibition scenarios, and other process upsets. If that level of analysis is not required, however, there are appropriate tools that permit reasonable steady state evaluation of digester systems.

**Steady State Models**

Many applications do not require the level of analysis provided by the dynamic models. Indeed, for most waste treatment processes, steady state or quasi-steady state behavior is critical to other dependent processes, and thus to overall plant performance. This is convenient for the designer, because steady state assumptions simplify mass balances considerably, but it is also a functionally accurate way to view many complex chemical processes (Levenspiel, 1999a).

Steady state models are essentially a mass balance on the total substrate. The first attempts at modeling of anaerobic processes utilized the first order and Monod kinetic models. While they do find place in the determination of kinetic coefficients for specific steps, they are not sufficient to capture steady state behavior in anaerobic digestion. Two of the most widely applied models for anaerobic digestion of high strength wastes, and particularly animal manures are Chen and Hashimoto’s (Chen et al., 1980; Hashimoto, 1982) adaptation of Contois’ method (1959), and Hill’s steady state approach as adapted from his dynamic model (1983). Husain (1998) offers a modification of Hill’s method that broadens its applicability to include a temperature function and to permit evaluation of any substrate based on the substrate biodegradability.

Hashimoto and Chen begin with a mass balance on the substrate and also incorporate a mass balance on the total bacterial population. They simplify the model inputs by lumping substrates and calculating availability as refractory COD, or bioavailable COD. This is convenient in working with complex substrates like manures where a significant portion of the waste is functionally indigestible. The basic model can be expressed as outlined in Eq 9:
\[
\frac{S_{T}}{S_{T_0}} = \frac{S + S_f}{S_{T_0}} = R + \frac{K \times (1 - R)}{\mu_m \times \theta - 1 + K}
\]  \hspace{1cm} (9)

Where:

- \(S_T\) = Effluent total substrate concentration
- \(S_{T_0}\) = Influent total substrate concentration
- \(S\) = Effluent biodegradable substrate concentration
- \(S_0\) = Influent biodegradable substrate concentration
- \(S_f\) = Nonbiodegradable material concentration
- \(R\) = \(\frac{S_T}{S_{T_0}}\)
- \(K\) = Dimensionless kinetic coefficient
- \(\mu_m\) = Maximum specific growth rate
- \(\theta\) = HRT

Biodegradable treatment efficiency (E) can also be determined via Equation 10:

\[
E = \left[ \frac{(S_0 - S)}{S_0} \right] \times 100
\]  \hspace{1cm} (10)

The principal valid criticism of the Chen-Hashimoto model is that it does not explicitly address volatile fatty acid (VFA) inhibition.

Hill's steady state model (1983) was developed based on simulation results from his dynamic model (1982). The model predicts methane production as a function of substrate loading rate. It does address inhibition broadly in that it predicts drops in methane production beyond certain stress states. Unfortunately, because of the way it is structured, the substrates that can be used are strictly defined lumped parameters, and the model as presented is only valid in the mesophilic range.

Husain developed another steady state solution to Hill's dynamic model by manipulating the equations so as to eliminate the time factor (viz. setting all derivatives with respect to time equal to zero). Equation 11 can then be iteratively solved for VFA:
\[
\frac{K_{dme}}{1 + \frac{K_{id}}{VFA}} + \frac{1}{\theta} = \frac{\mu_{mc}}{\frac{K_{sc}}{VFA} + 1 + \frac{VFA}{K_{sc}}}
\] (11)

Where:

- \(K_{dme} = K_{dm} = \mu_{mc} = \mu_m\) = maximum specific growth rate, 0.013T-0.129 (T=temp, °C)
- \(K_{idc}\) = Half velocity death rate constant for methanogens, 16 g VFA/L
- \(VFA\) = Volatile Fatty Acid concentration, g/L
- \(\theta\) = HRT
- \(K_{sc}\) = 3.0 g VFA/L

The VFA term can subsequently be plugged into Equation 12, which is then solved for S:

\[
\frac{K_{dm}}{1 + \frac{K_{id}}{VFA}} + \frac{1}{\theta} = \frac{\mu_m}{\frac{K_s}{S} + 1 + \frac{VFA}{K_i}}
\] (12)

Where:

- \(K_{id}\) = Half velocity death rate constant for acidogens, 16 g VFA/L
- \(K_s\) = 9.0 g Biodegradable Volatile Solids (BVS)/L
- \(S\) = Steady state substrate concentration, g/L
- \(K_i\) = VFA inhibition coefficient for acidogens, 11.0 g VFA/L

Equations 13 and 14 can be solved for \(\mu\) and \(\mu_c\), the specific growth rates of acidogenic and methanogenic consortia, respectively:

\[
\mu = \frac{\mu_m}{K_s + \frac{VFA}{S} + 1 + \frac{VFA}{K_i}}
\] (13)

\[
\mu_c = \frac{\mu_{mc}}{\frac{K_{sc}}{VFA} + 1 + \frac{VFA}{K_{sc}}}
\] (14)
Equation 15 can now be solved for $X$, the steady state concentration of acid forming bacteria, and

Equation 16 can subsequently be solved for $X_c$, the steady state concentration of methanogens:

$$\frac{S_0 - S}{\theta} + \frac{\mu X}{Y} = 0$$

(15)

$$\frac{VFA_0 - VFA}{\theta} + \frac{\mu X (1 - Y)}{Y} - \frac{\mu_c X_c}{Y_c} = 0$$

(16)

Where:

$Y = 0.1$ g organism/g BVS

$Y_c = 0.0315$ g organism/g VFA

To calculate methane production, solve Equation 17 for volumetric methane productivity ($1$ CH4/l reactor volume*d$^{-1}$):

$$VMP = 0.5 \times \left( \frac{\mu_c X_c (1 - Y_c)}{Y_c} \right)$$

(17)

Husain proceeded to apply the simplified Hill model to several datasets and demonstrate that the model provided reasonable predictions of steady state gas production value. The model requires valid estimates of influent biodegradability and VFA content for implementation, but may provide a reasonable compromise between the single step Chen and Hashimoto and the computationally intensive dynamic models outlined above.

**Temperature Effects**

Most anaerobic digestion takes place at mesophilic temperatures, between 30-38°C. Digesters can also be operated at higher thermophilic temperatures, typically 50-60°C. Thermophilic digestion is supposed to have several inherent advantages:

1. Assuming that stable biomass can be developed to function at high temperatures, the van’t Hoff-Arrhenius equation predicts that the product formation rate will approximately double
for every 10°C the temperature increases. The implication of this relationship is that required reactor volume is inversely proportional to temperature (Tchobanoglous et al., 2003).

2. Human pathogens are mesophiles. Exposure to thermophilic temperatures will disrupt cell walls and metabolic function and provide effective pathogen reduction.

3. Improved sludge dewatering characteristics (Parkin and Owen, 1986).

Potential disadvantages include higher energy requirements for heating, poor supernatant quality, and a less stable process.

Varel et al. (1977) generated a functional thermophilic seed culture from raw cattle waste over an eight day incubation period. He showed that in a semi batch process, he was able to generate and maintain a thermophilic culture capable of degrading a raw manure feed to produce biogas with approximately 50% methane content.

Subsequent work (Varel et al., 1980) considered reactors maintained at different temperatures from 30 to 60°C in 5-degree increments. The experiment showed that for extended HRTs there was no advantage to the higher temperatures. At lower retention times, however, there was a clear advantage in the higher temperatures with an approximately fourfold increase in biogas production for a reactor at 60°C than one at 40°C. It was also shown that higher temperature reactors could maintain proportionally higher substrate loading rates. While 60°C provided the best results, they were not sufficiently greater than those achieved at 50°C to justify the additional energy cost.

Mackie and Bryant (1995) also looked at fed batch thermophilic digestion of cattle waste. Their work showed higher energy content in the thermophilic biogas and higher reaction rates in the thermophilic systems. They also observed that the biological conversion efficiency of the mesophilic digester decreased by 49% between the lowest and highest loading rates, and by 16% for the thermophilic digester. Overall, they recommend using high VS loading rates and short HRT to capture the benefits of thermophilic anaerobic digestion.
REACTOR TYPES

Anaerobic digesters can be classified as reactors that retain biomass, and reactors that do not. The primary difference between these two types is that in a retained biomass reactor, the solids retention time (SRT) has been decoupled from the hydraulic retention time (HRT). A reactor can have CSTR hydraulic behavior with immobilized or attached growth solids or even CSTR with recycled solids resulting in the same effect (Buijs et al., 1982; Hulshoff Pol et al., 2004; Khakhar et al., 1999; Lomas et al., 1999; Michaud et al., 2002; Morgan, 1954; Schroepfer et al., 1955; Stander and Snyders, 1950; Taylor, 1972; Young and McCarty, 1969; Yu et al., 1999).

The microbial communities responsible for anaerobic digestion are complex in their interactions and slow growing. The slow growth rates directly impact process efficiency when treatment cultures are grown up from a starter culture. These issues can be addressed by maintaining the digestion populations separately from the waste material to be treated in an environment that is suited to their metabolic requirements. It has been shown that by doing so, populations can be evolved that will actually adapt to conditions that would generally be considered inhibitory (Parkin and Miller, 1982; van Velsen, 1977).

By decoupling HRT and SRT, several distinct advantages emerge. If the active solids are viewed as a catalyst for the anaerobic digestion reactions, the more concentrated the active mass, the more quickly the reactions can proceed. On the other hand, if the reaction rates are being controlled by the growth rates of the various populations in the reactor and their evolution as a syntropic community to the available substrate, this will inevitably control the overall rate of treatment as evidenced by the 20-40 day HRTs reported for batch systems started from seed cultures vs. the 4-12 hour HRTs for some high rate retained biomass digesters (Lettinga and Hulshoff Pol, 1991; Lettinga et al., 1980; Sung and Dague, 1995; Tchobanoglous et al., 2003). As a catalyst, the rate limiting mechanism in accomplishing the desired result is decoupled from the startup time required to develop the concentrations required to meet the treatment objective.

A second advantage is in adaptation. Provided that the feed is consistent, cultures will adapt to self optimize for the influent conditions. Different seed cultures will contain different organisms in different concentrations and if the seed culture is adapted to a feed that is different from the feed to be treated, it could require additional time for the culture to adapt before a stable community can develop. It is also
possible to build up cultures that are resistant to certain inhibitory compounds in the feed (Parkin and Miller, 1982; van Velsen, 1977). Finally, the higher concentration of active solids means higher substrate conversion rates, which correlates directly to shorter HRT for the same mass load and correspondingly smaller vessel requirements lower capital costs.

Having established that biomass retention is desirable, current practice utilizes both suspended and attached growth processes to bring this about. Suspended growth can include variations on CSTR with sludge recycle where active solids are removed from the effluent and returned to the digester (Morgan, 1954; Stander and Snyders, 1950), and self-selecting processes like sludge blanket and anaerobic sequencing batch reactors where biomass granulates or flocculates to become denser than the surrounding water and settles to the bottom of the vessel (Hulshoff Pol et al., 2004; Lettinga et al., 1980). The effluent is removed as supernatant, retaining the denser active solids in the reactor. There are also attached growth processes where anaerobic biomass forms a biofilm on a media surface and feed is passed over the biofilm to provide the contact necessary for removal (Michaud et al., 2002; Taylor, 1972).

In all cases, however the active biomass is retained in or returned to the reactor while the fluid containing the substrate passes through. The biomass effectively acts as a retained catalyst while the substrate in the feed completes its reaction as it transits the reactor volume. The balance between the reaction rate, the contacting pattern of the feed and catalyst, the catalyst concentration, and the time that the average substrate particle spends in the reactor determines the efficiency of the process.

The Induced Bed Reactor

The reactors under consideration in this research program were developed from the principles of the Upflow Anaerobic Sludge Blanket (UASB) reactors originated by Dr. Gatze Lettinga (Figure 1-7) (Lettinga and Hulshoff Pol, 1991; Lettinga et al., 1980). His research group observed that under certain conditions, anaerobic consortia would form granular particles (Hulshoff Pol et al., 2004). These particles settled easily in aqueous solutions and formed relatively dense beds in the bottoms of vertical upflow bioreactors. They developed an operational strategy that provided substrate to the active granulated solids and showed that this process design yielded very effective reactor designs with removal efficiencies as high as 95% over HRTs as low as 4-12 hours.
Figure 1-7: Cutaway view of an Upflow Anaerobic Sludge Blanket (UASB) reactor.

Reactor configuration is such that influent enters from the bottom via a diffuser to provide equal distribution through the sludge blanket and minimize short-circuiting. Substrate transits the sludge blanket vertically, converting COD to biogas and some biomass. Biogas is directed via the deflector baffles to the 3-phase separator, while treated liquid effluent flows around the baffles and the separator and over a weir to the reactor outlet. UASBs operate on the principle of separation of solids and hydraulic retention times (SRT and HRT) via granulated biomass retention; the microbial consortia required for anaerobic digestion are slow to reproduce, but if they can be retained in a reactor while the fluid carrying the dissolved carbonaceous substrate that they feed on can pass freely through it at a rate that optimizes the metabolic ability of the active solids to convert dissolved solids to biogas, the volumetric efficiency of the reactor increases substantially. The active biomass acts as an effective filter, metabolizing available
carbon to CO₂ and CH₄. This strategy is most effective when the available carbon is well dissolved so as to optimize the difference in physical properties between the solids to be retained in the reactor and the liquid that is to be passed through.

In 1993, Dr. Conly Hansen of Utah State University began working with digestion of high-suspended solids, high strength wastewaters in reactors configured as UASBs in an attempt to translate some of the advantages of the high-rate small footprint systems to agricultural and food wastes. This resulted in the development of the Induced Bed Reactor (IBR) (Figure 1-8), an upflow design that preserves some of the biomass concentrating characteristics of the UASB while allowing for the treatment of complex high solids and wastewaters such as dairy manures.

The principal problems addressed by the IBR that make it unique from other upflow digesters are in the design of the influent and phase separation portions of the digester and in the operational strategy. There is also recognition that as a retained biomass reactor receiving relatively large quantities of functionally inert solids, the IBR cannot have the same solids retention effectiveness as the UASB without accumulating solids and plugging.

The system was designed around dairy manure as the controlling fluid. Manure is a complex waste product with widely ranging characteristics depending on temperature, feed, collection mechanism, age, water content, and other environmental factors (Cheong and Hansen, 2006; El-Mashad et al., 2005; Garcia-Ochoa et al., 1999; Janzen, 1999; Karim et al., 2007). The fluid properties control the inlet and outlet conditions and the phase separator design. In order to accommodate the high solids content and large maximum particle sizes anticipated in the influent, a relatively large inlet is required to permit the solids to pass without plugging the feed line. In order to maintain sufficient velocity in the feed line to prevent plugging and still maintain the capacity to feed semi-continuously, the distributed feed of the UASB had to be reconfigured to a single feed. A diffuser plate was added to the digester to provide some distribution of the influent into the sludge blanket. The reactor is tall (10 m) relative to typical UASB designs (5-8 m) (Lettinga and Hulsloot Pol, 1991) to provide additional separation time for active solids to settle in the reactor. Current design practice maintains reactor diameter as a function of height with an aspect ratio of 2.5, although the importance of this criterion on digester performance has not been demonstrated.
Figure 1-8: Cutaway view of Induced Bed Reactor (IBR).

In the top portion of the reactor, the IBR channels the effluent flow through a relatively small hole in the septum (phase separator), whereas the UASB provides a large cross sectional area relative to the area of the separator. The IBR configuration permits the concentration of floatables that could potentially plug the reactor in a zone where a mechanical auger can either lift them into the upper chamber, or push them back down into the main reactor volume depending on operational requirements.

Finally, the outlet configurations are different. The UASB uses a weir configuration to provide an outlet condition with large area and therefore minimal flow velocity to help retain solids and return them to the digester. The IBR utilizes an inverted trap with a much smaller area to maintain the desired gas pressure in the unit and to provide a means of passing large solids that may accumulate.
Operationally, they are different as well. The IBR is typically operated at HRTs of 3-7 days with waste strengths of 25-85 kg/m² COD with 30-60% reduction, while the UASB is operated at 4-14 hours with waste strengths of 1-18 kg/m² COD with 90-95% reduction. The UASB is capable of treating both high and low strength wastes effectively and to a high degree, but the substrates must be relatively clean and free of non-volatile suspended solids that could displace or interfere with active solids (Lettinga and Hulschoff Pol, 1991; Lettinga et al., 1980).

The principal theoretical advantages of the IBR are that it a) reduces the need for solids separation over what would be required for other high-rate systems, b) permits the treatment of all the material in the feed stream, and c) with a 4-day HRT requires 5x less volume than a similarly performing plug or CSTR digester with a 20-d HRT. The IBR was developed specifically to treat high strength, high solids wastes. The physical constraints imposed by these objectives dictate an inherently less efficient process in terms of overall process efficiency than other high rate systems, but one that nevertheless may have a niche in treatment technology.

REFERENCES


CHAPTER 2
THE INDUCED BED REACTOR ANAEROBIC DIGESTER

Anaerobic Digestion utilizes naturally occurring microbial consortia to anaerobically degrade substrates to high energy value biogas. Over the past century, researchers have worked to capture and optimize the benefits of anaerobic digestion and mitigate the difficulties. One of the most significant advances in this work was the development of high rate retained biomass anaerobic reactors enabling the relatively rapid stabilization of high strength wastes.

The Induced Bed Reactor (IBR) was developed at Utah State University (USU) to apply high-rate anaerobic digestion techniques to highly-suspended solids content substrates (6-12% total solids) such as food waste and dairy manures. The IBR is a vertical upflow retained mass bioreactor similar to the Upflow Anaerobic Sludge Blanket (UASB) which inspired the original design, but with key differences to enable the reactor to operate at high solids concentrations.

This technology has been successfully implemented at full-scale multiple installations in the United States and Canada for treatment and energy production. Installations range in size from 135 to 3000 milking head with throughputs of 27-300 m$^3$·d$^{-1}$. The reactor appears to be capable of treating relatively high solids substrates such as the dairy manures considered herein with calculated influent TS concentrations ranging from 3.9-10.4%. It appears that relatively higher solids loading rates can be correlated to higher specific biogas production without reactor failure for the conditions observed.

INTRODUCTION

The purpose of anaerobic digestion in wastewater treatment is to reduce waste sludge volume and activity by fermenting the waste in the absence of oxygen. Anaerobic digestion relies on the symbiotic relationship between two general classes of anaerobic microorganisms to catabolize carbonaceous substrates to the relatively stable and innocuous end products of biosolids and biogas (Bryant et al., 1967).

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1 Coauthored by J. S. Dustin, J. D. Dustin, and C. L. Hansen
The relatively long doubling times of anaerobes require either long residence times (20-40 d) to allow populations to accumulate to levels permitting significant treatment or separation of hydraulic retention time (HRT) and solids residence time (SRT) to permit high-rate treatment by retaining the active microbial solids in the treatment volume. Methods for accomplishing this include separating the active solids from the effluent and recycling them to the digester, by directly utilizing an attached growth process, or through the use of an induced bed or blanket of solids that self-select to remain in the reactor as the substrate passes through (Gerardi, 2003; McCarty, 1981; Tchobanoglous et al., 2003). High rate processes like these generally require a relatively low suspended solids concentration in the influent to avoid dilution of the active biomass and operational problems in the digester (plugging, sedimentation, short-circuiting, etc).

The inability to treat wastes with high influent suspended solids content poses a problem application of high rate processes to substrates like those associated with food processing wastes and livestock manures. In the early 1990s, Dr. Conly Hansen at Utah State University proposed a digester configuration, the Induced Bed Reactor (IBR) (Hansen and Hansen, 2005) that was designed to address the operational limitations of applying high rate treatment techniques to high strength, high solids substrates.

The IBR relies on biomass retention in the digester to provide an accumulation of active solids, similar to the operational principles first proposed by Dr. Gatze Lettinga (1980) for his Upflow Anaerobic Sludge Blanket (UASB) reactor design (Figure 2-1). UASBs operate on the principle of separation of SRT and HRT via granulated biomass retention. The active biomass acts as an effective filter, metabolizing available carbon to CO₂ and CH₄. In the UASB, influent enters the reactor from the bottom via a diffuser to provide equal distribution through the sludge blanket and minimize short-circuiting. Substrate transits the sludge blanket vertically, converting chemical oxygen demand (COD) to biogas. Biogas is directed via the deflector baffles to the 3-phase separator, while treated liquid effluent flows around the baffles and the separator and over a weir to the reactor outlet.

In 1993, Dr. Hansen began working with digestion of high-suspended solids, and high strength wastewaters in reactors configured as UASBs in an attempt to translate some of the advantages of the high-rate small footprint systems to agricultural and food wastes resulting in the development of the Induced Bed Reactor (IBR) (Figure 2-1), an upflow design that preserves some of the biomass concentrating
characteristics of the UASB while allowing for the processing of complex high solids and high strength wastewaters such as dairy manures.

Figure 2-1: Cutaway view of an Upflow Anaerobic Sludge Blanket (UASB) reactor (L) and an Induced Bed Reactor, IBR (R).

The IBR is designed as a modular system where reactor vessels are generally shop fabricated from steel plate and transported to the site for placement. Shipping considerations dictate a maximum diameter of 4.1 m (13.5 ft). The height:diameter aspect ratio of the digester is generally maintained at approximately 2.5:1, yielding an overall reactor height of 9.8 m (32 ft) at full scale. Wall thickness, internal reinforcement, and anchorage are dictated by building code requirements.

Influent enters the reactor at the base via a 5 cm (2 in) minimum diameter steel pipe. Pipe size is determined by flow rate, size of solids in the system, potential of solids to bridge the pipe diameter, and the need to maintain a minimum cleansing velocity in the line. Experience has shown that a 5 cm (2 in) minimum diameter steel pipe is appropriate, and 5 cm/s (2-ft/s) is the minimum recommended line velocity (Illinois et al., 2004). The influent pipe is routed to the center of the reactor where a tee is placed to direct
the influent into two split streams, one pointing down at the tank floor, the other pointing up at a steel diffuser plate.

The diffuser plate occupies one half of the reactor area and is located approximately 30 cm (12 in) from the tank floor. There is a baffle (sometimes called a septum) located at approximately 90% of the reactor height. The septum is a steel cone with a mild slope where the peak is removed to leave an opening of 10-30 cm (4-12 in). The purpose of the upper baffle/septum is to provide a contact surface to encourage separation of the solids and biogas bubbles, and to provide a restriction where flow velocity increases as a function of cross sectional area and flow rate, transporting solids out of the lower volume. There is an auger located in the center of the septum/baffle hole rotating at 0.5-1 RPM to keep the septum hole clear of bridging solids. The auger shaft also supports a bar which Rotates about the central axis of the reactor, constantly sweeping the top of the wet volume to discourage foam and crust formation.

Liquid reactor effluent exits via a submerged 25 cm (10 in) diameter outlet in the upper reactor volume. Biogas exits the top of the reactor, generally via a 5 cm (2 in) diameter line. Pressure in the reactor is maintained via two water columns, one on the liquid effluent line and one on the gas effluent line. Liquid effluent exits via an inverted trap and overflows a circular weir, generally to an open channel and thence to receiving lagoons. Gas bubbles out through a submerged outlet and is routed for further use. The liquid effluent water column height is fixed by the hard piping of the reactor effluent drain. The gas effluent water column height can be adjusted by raising or lowering the water level in the gas effluent trap. Full scale IBRs typically operate at 20-36 cm (8-14 in) of water pressure.

The differences between IBRs and UASBs are derived from the need to address the high solids content of the influent. As previously noted, the influent of the IBR is typically a single line double jet directed at a diffuser plate and bottom of the tank to provide some diffusion of the influent into the sludge bed. This configuration is dictated by the high solids content of the influent and the need to prevent the solids from settling in and plugging the influent line.

The UASB runs at higher flow rates and relies on a relatively low suspended solids influent. UASB feed piping design criteria therefore allows design of influent distribution systems where line sizes are less restricted, and a more optimal feed distribution can be achieved to ensure that the influent is well
distributed through the cross sectional area of the digester as it migrates up through the sludge blanket (Lettinga and Hulschhoff Pol, 1991; Lettinga et al., 1980).

Exit conditions are also fundamentally different. UASB effluent exits the reactor via an overflow weir designed to provide a large surface area relative to the volumetric flow rate and thus reduces effluent velocity and enhances solids retention at the exit point. In order to minimize plugging in the effluent line, the IBR utilizes a submerged outlet of a relatively small cross sectional area to permit higher velocities and carry suspended solids out of the system.

The principal advantages of the IBR system are that the system a) reduces the need for solids separation over what would be required for other high-rate systems, b) permits the processing of all the material in the feed stream, and c) with a 4-day HRT requires five times less volume than a similarly performing plug or complete mix digester with a 20-d HRT.

MATERIALS AND METHODS

In order to evaluate existing IBR systems, data was solicited from system operators regarding location, startup date, number of reactor tanks in the systems, number of cattle feeding waste into the systems, substrate collection methods, number of reactor tanks, and gas and influent flow rates. Hydraulic retention time (HRT) was calculated by dividing total reactor volume by influent flow rate. A mass balance was then used to calculate influent loading conditions for COD and total solids (TS), and to calculate a theoretical COD removal based on reported biogas flow rates.

A mass balance based on the schematic presented in Figure 2-2 was constructed on COD for each system (Equation 1). COD can be calculated based on the stoichiometry of the reactions required to completely oxidize reactor contents, and therefore provides a reliable common denominator for use in quantifying mass transfer through the system.

Influent COD (COD_{in}) was calculated based on tabulated values presented in ASABE D384.1 (ASABE, 2003) of 11 kg COD·(1000 kg live animal mass·d)^{-1}, 640 kg live animal mass·(live animal)^{-1} and the number of cows reported for each system. COD_{bio} was calculated given that rate of biomass accumulation in a reactor at steady state is, by definition, equal to zero (dCOD_{bio}/dt = 0). Biogas quality (bq), the unitless ratio of volume of CH₄ to volume of biogas was assumed to be 0.65, consistent with
average values reported in the literature (Gerardi, 2003; Tchobanoglous et al., 2003). Specific biogas production rate \( \text{sbpr} = \text{L biogas produced} \cdot (\text{L wetted volume of reactor} \cdot \text{d})^{-1} \) was calculated from the reported biogas production rate. The mass balance was then solved for COD\(_{BG} \) using Equation 2.

![Diagram of mass balance model of Induced Bed Reactor](image)

**Figure 2-2:** Schematic for mass balance model of Induced Bed Reactor. COD = chemical oxygen demand; \( X \) = suspended solids concentration

\[
\text{COD}_{\text{INF}} - (\text{COD}_{\text{LE}} + \text{COD}_{\text{RI}}) = \frac{d\text{COD}_{\text{RET}}}{dt} \quad (1)
\]

\[
\text{COD}_{\text{BG}} = \frac{\text{sbpr} \cdot 0.378 \text{L CH}_4}{\text{g COD}} \quad (2)
\]

**RESULTS AND DISCUSSION**

Data for IBR systems at five locations in the Northern US and Ontario, Canada are presented in Table 2-1. The data are self-reported and must therefore be regarded with some caution. There are too few data points to be statistically significant or to draw substantive conclusions regarding treatment performance, but it is interesting to note the differences in operational parameters and how those differences appear to impact system performance when evaluated using equivalent assumptions regarding biogas quality and manure characteristics as previously outlined.

Data is presented for dairies ranging from very small installations to large corporate operations. The highest specific biogas production rates were calculated for the facilities with the lowest hydraulic retention times and highest organic loading rates, indicating that there may be no benefit for longer HRTs in the
systems if maximization of volumetric efficiency for biogas production is a significant design criteria. This observation adds at least anecdotal strength to the argument that the IBR possesses characteristics favorable to the processing of high suspended solids waste and demonstrates that the IBR can operate well outside the UASB parameters in terms of solids loading.

Table 2-1: Summary of reported and calculated characteristics for Induced Bed Reactor anaerobic digester facilities

<table>
<thead>
<tr>
<th>Location</th>
<th>Jer-Lindy Farm</th>
<th>Huls Dairy</th>
<th>Stanton Dairy</th>
<th>Wadeland Dairy</th>
<th>Whitesides Dairy</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Brooten, MN</td>
<td>Corvallis, MT</td>
<td>Ontario, Canada</td>
<td>Ogden, UT</td>
<td>Minnedoka, ID</td>
</tr>
<tr>
<td>Startup Date¹</td>
<td>06/2008</td>
<td>11/2008</td>
<td>09/2008</td>
<td>10/2004</td>
<td>04/2005</td>
</tr>
<tr>
<td>Number of Cows¹</td>
<td>135</td>
<td>340</td>
<td>800</td>
<td>1000</td>
<td>3000</td>
</tr>
<tr>
<td>Substrate Collection Method¹</td>
<td>Scrape/Flush</td>
<td>Scrape/Flush</td>
<td>Scrape/Flush</td>
<td>Scrape/Flush</td>
<td>Scrape/Vacuum</td>
</tr>
<tr>
<td>Number of Reactor Tanks¹</td>
<td>1</td>
<td>2</td>
<td>8</td>
<td>3.6²</td>
<td>10</td>
</tr>
<tr>
<td>Total Reactor Volume (m³)¹</td>
<td>114</td>
<td>227</td>
<td>909</td>
<td>409</td>
<td>1,136</td>
</tr>
<tr>
<td>Influent Flow Rate (m³·d⁻¹)¹</td>
<td>26.9</td>
<td>54.9</td>
<td>122</td>
<td>74.2</td>
<td>303</td>
</tr>
<tr>
<td>Operating Temp (°C)¹</td>
<td>39</td>
<td>38</td>
<td>41</td>
<td>37</td>
<td>38</td>
</tr>
<tr>
<td>Calculated HRT (days)</td>
<td>4.2</td>
<td>4.1</td>
<td>7.5</td>
<td>5.5</td>
<td>3.8</td>
</tr>
<tr>
<td>Influent OLR (kg COD·m⁻³·reactor vol·d⁻¹)</td>
<td>7.9</td>
<td>10.0</td>
<td>5.9</td>
<td>16.3</td>
<td>17.6</td>
</tr>
<tr>
<td>Specific Biogas Production (m³ biogas·m⁻³ reactor vol·d⁻¹)</td>
<td>2.7</td>
<td>3.4</td>
<td>2.3</td>
<td>3.2</td>
<td>3.8</td>
</tr>
<tr>
<td>Specific CH₄ Production (m³ CH₄·m⁻³ reactor vol·d⁻¹)</td>
<td>1.7</td>
<td>2.2</td>
<td>1.5</td>
<td>2.1</td>
<td>2.5</td>
</tr>
<tr>
<td>Specific COD removal as CH₄ (kg COD·m⁻³ reactor vol·d⁻¹)</td>
<td>4.6</td>
<td>5.9</td>
<td>4.0</td>
<td>5.5</td>
<td>6.5</td>
</tr>
<tr>
<td>Influent TS (%)</td>
<td>3.9%</td>
<td>4.8%</td>
<td>5.0%</td>
<td>10.4%</td>
<td>7.6%</td>
</tr>
<tr>
<td>Influent TS Loading Rate (kg COD·m⁻³·reactor vol·d⁻¹)</td>
<td>10.9</td>
<td>8.6</td>
<td>6.4</td>
<td>17.8</td>
<td>19.2</td>
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</tbody>
</table>

¹ Self-reported values as transmitted to the authors by Andigen, LLC
² Wade Dairy utilizes a 60% volume reactor in addition to three full size reactors

It is also interesting to note that the smaller installations consistently show significantly lower influent TS as % and as loading rates than the larger installations. The authors have observed that feed pumps at the installations vary, with the three smaller dairies utilizing submersible centrifugal pumps and the larger installations using rotary lobe (Wadeland) and progressive cavity (Whitesides) displacement pumps. The displacement pumps are inherently capable of pumping the heavier slurries characteristic of the higher
solids loadings; this gives operators the opportunity to treat significantly more waste solids per reactor volume than a system that is limited by the solids handling capacity of the feed pumps given that for a given unit solids loading, the volume occupied by a 5% solids slurry will be twice that occupied by a 10% slurry. When costs like heating of influent to reactor temperature and pumping power are factored in, the additional expense of displacement pumps may be justifiable.

CONCLUSIONS

The IBR has been implemented in the United States and Canada at the five facilities reviewed with reasonable treatment success. The reactor appears to be capable of treating relatively high solids substrates such as the dairy manures considered herein with calculated influent TS concentrations ranging from 3.9-10.4%. It appears that relatively higher solids loading rates can be correlated to higher specific biogas production without reactor failure for the conditions observed, although it must be recognized that observations reported herein are derived from self-reported data, and should therefore be regarded with caution.

It is hoped that this information will serve as an aid to future designers in making decisions regarding implementation of IBR systems, and to future researchers in identifying questions that will help to further understanding of IBR performance. Future work should include evaluation of the IBR to determine contacting patterns and mixing behavior, the ability of the reactor to retain solids, and long term monitoring of plant scale reactors to develop an objective understanding of IBR performance.

REFERENCES


CHAPTER 3
HYDRODYNAMIC MODELING OF THE
INDUCED BED REACTOR ANAEROBIC DIG ESTER

The Induced Bed Reactor (IBR) was developed at Utah State University (USU) to apply high-rate anaerobic digestion techniques to highly-suspended solids content substrates (6-12% total solids). This technology has been successfully implemented at multiple full-scale installations in the United States and Canada as a waste treatment and energy production technology. Residence Time Distribution (RTD) studies for 58 L lab-scale reactors operated at a 3.8-d hydraulic retention time were conducted at three temperatures (35°C, 45°C, and 55°C) under both control (no active biomass, no reaction taking place) and active digestion conditions. Rhodamine WT and Li+ were used as tracers. Rhodamine appears to interact with the digester contents, raising questions about its suitability as a tracer in this context.

The results show that the IBR most closely approximates Completely Stirred Tank Reactor (CSTR) behavior when operated under the study conditions. A compartment real CSTR model, incorporating elements of dead zone and bypass flow appears to be the most appropriate representation of the data. Mixing is likely due to a combination of energy inputs from thermal gradients induced by heat flux through the reactors and reactor and shear rates induced by gas evolution in the sludge bed.

INTRODUCTION

The purpose of anaerobic digestion in wastewater treatment is to reduce waste sludge volume and activity by fermenting the waste in the absence of oxygen. The process relies on the symbiotic relationship between two general classes of anaerobic microorganisms to catabolize carbonaceous substrates to the relatively stable and innocuous end products of biosolids and biogas (Bryant et al., 1967). Process designers must provide an environment that will permit efficient distribution of substrate and biomass and provide optimal conditions for the required reactions to take place. An understanding of process microbiology and ecology is obviously essential to this effort (Chen et al., 1980; Hill, 1982, 1983; Husain,

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1 Coauthored by J. S. Dustin and C. L. Hansen
1998; Speece et al., 2005), but of no less importance is an understanding of the hydrodynamic behavior of the reactor under consideration (Levenspiel, 1999a).

As early as 1935 it was observed that tracers could be used to construct curves of concentration vs. time that provided insights into mixing behavior in reactors (MacMullin and Weber). Danckwerts, in his seminal 1953 paper, formalized an approach wherein the extremes of mixing possibilities (viz. plug (piston) flow and completely mixed vessels) were defined and then proceeded to apply principles of statistical analysis to normalized tracer responses curves, normalizing the curves and treating them as statistical distributions (residence time distributions, RTDs) and demonstrated the relationships between response curves and the implications of deviations in real data from the ideals of plug flow and complete mix. Subsequent researchers expanded on this work to develop specific modeling strategies that could then be incorporated with kinetic models to provide mass-balance based comprehensive models of reactor behavior (Fogler, 2006; Levenspiel, 1993, 1999b; Wen and Fan, 1975).

The Induced Bed Reactor (IBR) (Figure 3-1) was developed by Dr. Hansen at Utah State University as an adaptation of immobilized biomass reactor technology to permit treatment of high strength, high solids substrates. The reactor was developed and scaled to accommodate the unique handling requirements of high solids slurries such as dairy manures and food processing wastes. Currently there are approximately 4,000 m³ of installed capacity in the US and Canada.

The IBR may be classified as an immobilized biomass upflow reactor. It operates similarly to the more widely known UASB with some important distinctions as outlined in Chapter 2. The reactors are typically operated in the mesophilic temperature range between 28° and 35°C. Substrate is generally high solids (4-10% TS) dairy manure or food waste. The sludge bed is formed by the gradual development of a self-selecting granular sludge and maintained by the in-situ segregation of the more dense sludge particles and the liquid fraction of the reactor contents.

The reactors are operated as a continuous process. Influent enters the reactor from the bottom as a pulsed feed (on/off cycles are generally required to maintain sufficient velocity in the feed lines to maintain solids in suspension and meet target hydraulic retention time, HRT) where it is dispersed into the reactor volume via contact with a diffuser plate. Substrate transits the sludge bed and liquid volume, exiting via the
submerged outlet and inverted trap. Biogas is constantly evolving as a result of the biochemical reactions between the influent substrate and the sludge bed. It exits via the gas outlet at the top of the vessel. The function of the septum is to intercept sludge particles attached to gas bubbles and provide an opportunity for separation and retention of the sludge. The auger keeps the hole in the septum clear. The foam bar at the liquid/gas interface serves as a disruptor to knock down foam and prevent it from filling the headspace.

Figure 3-1: Cutaway view of Induced Bed Reactor (IBR).

RESIDENCE TIME DISTRIBUTION (RTD)

The purpose of this research was to determine the residence time distribution (RTD) of the IBR which represents the variation in residence time experienced by matter flowing through the system. This information permits direct observation of the amount of time that reactants spend in a vessel, which in turn provides input parameters for the mass balance model. The curves developed from an RTD study can also be used to determine the volumetric efficiency of the system by identifying and quantifying bypassing, channeling and dead space (Danckwerts, 1953; Fogler, 2006; Levenspiel, 1993; Wen and Fan, 1975).

RTD is determined by injecting a known mass of an appropriate tracer (Denbigh and Turner, 1965) at the reactor inlet with the attenuated concentration being observed at points of interest in the reactor. The
two primary experimental methods involve injection of the tracer as a slug (instantaneous injection at the inlet) or as a step function (tracer is introduced at a constant concentration over the duration of the experiment). The concentration of the tracer is plotted against time (C-T curve). If concentration is normalized to the maximum input concentration and time is normalized to the idealized hydraulic retention time, the curve that can be derived from the normalized values is the E-curve for a pulse input and the F-curve for a step input. The E-curve represents the residence time distribution of the material in the reactor and the F-curve represents the cumulative residence time distribution. If either is known, the other can be derived given that the E-curve function is the derivative of the F-curve function.

**Ideal Reactors**

The two ideal reactor types for steady state conditions are the completely stirred tank reactor (CSTR), and the plug flow (PF) reactor. For any given reaction, one of these ideals will provide the most efficient conversion scenario. Plug flow is the condition where elements of a fluid entering a vessel at the same time move through it with constant and equal velocity on parallel paths and leave at the same moment. In an ideal CSTR, the contents are completely mixed such that the properties are uniform throughout the reactor. Mathematically, the distribution of tracer residence time in the system for a CSTR can be represented by an exponential decay function (Equation 1) where \( C_t = \text{tracer concentration at time } t \), and \( t = \text{sample time} \):

\[
C_t = 1 - e^{-t}
\]

(1)

**Non-Ideal Reactors**

In reality, no reactor will meet these ideals perfectly and may incorporate elements of PF, CSTR, dead zones, and bypassing in varying quantities and configurations. Wen and Fan (1975) summarize many of the real system models and provide methods of analysis to evaluate RTDs and quantify reactor behavior based on tracer study results.

**Materials and Methods**

**Experimental Apparatus**

The experimental apparatus is shown in Figure 3-2. Three 58 L working volume IBRs were constructed
from extruded acrylic tubing 0.30 m ID and 0.91 m long (Figure 3-3). Digesters were fed from a common 800 L feed tank. The feed tank was maintained in a cold room at 5°C (±1.5°C) to minimize unwanted microbial growth and mixed using a Lightnin variable speed mixer (EV1P25M1C48, Lightnin, Rochester, NY). Influent feed was controlled by a programmable peristaltic pump (Masterflex Cole-Parmer, Inc., Vernon Hills, IL) with stacking heads. Feed tubing was randomly rotated between heads to minimize feed rate variations. During the experiments, the reactors were maintained at 35°C, 45°C, and 55°C (±1.5°C) respectively by heat tape applied to the base of the reactors and controlled by Love TS-13011 digital temperature switches.

![Diagram](image)

**Figure 3-2: Schematic design of research apparatus.**

Effluent gas was routed through a liquid capture foam trap, thence to refrigerated water columns where excess water vapor was removed by bubbling the gas through the columns at 5°C before warming the gas to 22°C for flow measurement. Gas samples were collected in Tedlar bags (Fisher Scientific) for composition analysis, with excess gas vented via a fume hood. Effluent samples were collected for analysis in 50 mL centrifuge tubes. For the tracer studies, all effluent was collected in polyethylene containers to permit composite sample analysis. Foam control was provided by a semi-continuous injection of Sigma Antifoam B (Sigma Aldrich, Inc. St Louis, MO) into the reactor headspace via LMI AA97 series metering pumps (Milton Roy, Ivyland, PA) to maintain a headspace antifoam concentration of 30 mg/L.
INSTRUMENTATION AND DATA COLLECTION

Gas flow rates were monitored using Sierra Instruments thermal mass flowmeters (Monterrey, CA, 822S-L-2-ON1-PV1-V4). Temperatures and gas flow rates were logged continuously using Labview 8.2 (National Instruments, Austin, TX) running on an IBM clone PC with Windows XP operating system and National Instruments (NI) hardware. NI data acquisition hardware included a cDAQ-9172 USB Chassis, and NI 9203 and 9211 modules. pH measurements were recorded from daily grab samples using a Ross Ultra electrode and a Thermo Scientific Orion 720A+ meter (Thermo Scientific, Pittsburgh, PA).

EXPERIMENTAL DESIGN

Twelve studies were conducted in the three reactors including a control study in clean water and a second study which incorporated the effects of active biomass and digestion on mixing in the systems at steady state (Table 3-1). Tracers were mixed into the digester feed tank at known concentrations, and fed to the reactors via the influent feed pump as a step input. Effluent samples were collected from the effluent line at recorded time intervals and comprised both mixing cup and grab samples. Mixing cup samples were
incorporated into the grab sample dataset by collecting the full volume of the effluent between grab sample intervals, stirring the sample, taking a representative aliquot for analysis, and recording the calculated average time corresponding to the sample interval. The structure of the sampling program permits the assumption of closed boundary conditions at the entry and exit.

<table>
<thead>
<tr>
<th>Description</th>
<th>Clean Water</th>
<th>Active Digestion 1</th>
<th>Active Digestion 2</th>
<th>Active Digestion Washout</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tracer</td>
<td>RWT&lt;sup&gt;a&lt;/sup&gt;</td>
<td>RWT&lt;sup&gt;a&lt;/sup&gt;</td>
<td>Li&lt;sup&gt;b&lt;/sup&gt;</td>
<td>RWT&lt;sup&gt;a&lt;/sup&gt;</td>
</tr>
<tr>
<td>[Tracer]&lt;sup&gt;a&lt;/sup&gt;, mg·L&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>1.5</td>
<td>1.5</td>
<td>1.0</td>
<td>10</td>
</tr>
<tr>
<td>COD&lt;sub&gt;int&lt;/sub&gt;, mg·L&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>--</td>
<td>30.6 (2.1)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>OLR&lt;sub&gt;int&lt;/sub&gt; gCOD·L&lt;sup&gt;-1&lt;/sup&gt;·d&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>--</td>
<td>8.05 (0.9)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total Solids&lt;sub&gt;int&lt;/sub&gt;, % (σ)</td>
<td>--</td>
<td>2.16 (0.27)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fixed Solids&lt;sub&gt;int&lt;/sub&gt;, % (σ)</td>
<td>--</td>
<td>0.88 (0.15)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatile Solids&lt;sub&gt;int&lt;/sub&gt;, % (σ)</td>
<td>--</td>
<td>1.28 (0.27)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Temperature Levels (°C)</td>
<td>35, 45, 55</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>HRT (days)</td>
<td>3.8</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<sup>a</sup> RWT = Rhodamine WT (Kingscote Chemicals, Miamisburg, OH)
<sup>b</sup> Li+ = Lithium ion from Lithium Chloride (Sigma Aldrich, St Louis, MO)

**Tracer**

Tracers selected were Rhodamine WT (RWT) (Kingscote Chemicals, Miamisburg, OH) and lithium ion from lithium chloride (Li+) (Sigma Aldrich, St Louis, MO). A confirmatory study was performed on the lab scale reactors using Li+ as lithium chloride to verify the tracer mass balance. The study design does not permit quantitative evaluation of tracer performance in the effluent matrix. Rhodamine WT is known to have two isomers, one of which exhibits sorptive behaviors in certain environments (Vasudevan et al., 2001). Lithium also has the disadvantage of being inhibitory in anaerobic systems at concentrations in excess of 250 mg/L (Anderson et al., 1991), although this is primarily a concern in pulse input studies where very high concentrations of tracer are anticipated at the injection site before mixing distributes the tracer in the reactor volume. These effects are confounded with other variables in the study and cannot be separated, but both chemicals are widely regarded as appropriate for digester RTD modeling in spite of their limitations (Denbigh and Turner, 1965; Leighton and Forster, 1996; Lou et al., 2006; Tchobanoglous et al., 2003).
It proved difficult to calibrate the RWT fluorescence to concentration consistently; the fluorescence response, while linear, varied with reactor temperature. The physical experimental design (viz. selection of the step input function) does permit two point calibration for each treatment which is validated by the linear response of the various RWT calibration curves. This approach requires that the calibration be self referencing to an observed minimum and maximum for each sample set. The result is that while the RWT studies are acceptable in determining mixing patterns, they cannot be used to reliably determine compartment volumes in the compartment model (Fogler, 2006; Levenspiel, 1999b).

**Medium Composition**

The substrate solution had a COD of 30,600 mg/L, composed of 23,438 mg·L⁻¹ of dextrose, plus the following nutrients (mg·L⁻¹): 2,500 yeast extract, 2,656 NH₄Cl, 525 K₂HPO₄, 225 FeCl₂ _ 4H₂O, 469 CaCl₂ _ 2H₂O, 391 MgSO₄ _ 7H₂O, and 313 KCl. To prevent required microbial trace element deficiency, a trace nutrient solution comprised of (mg·L⁻¹) 500 H₂BO₃, 500 ZnCl₂, 300 CuCl₂, 500 MnSO₄ _ H₂O, 500 (NH₄)₆Mo₇O₂⁴ _ 4H₂O, 500 AlCl₃, 500 CoCl₂ _ 6H₂O, and 500 NiCl₂ was added by 0.01% (v/v) to each nutrient medium batch. 10,000 mg·L⁻¹ NaHCO₃ was added to maintain initial buffering capacity, and tap water (City of Logan, UT) was used as dilution water. The components were similar to those used for cultivating anaerobic bacteria in a high rate suspended biomass reactor by Cheong and Hansen (2008).

**Culture Development**

Cultures were developed from seed sludge taken from an operating IBR utilizing a dairy manure feed (Wade Dairy, Ogden, UT). The seed sludge was passed through a ¼” mesh screen to remove larger solids. The reactors were filled with a mixture of 1/3 sludge, 1/3 water, and 1/3 medium, and brought to temperature over 5 days. The reactors were then allowed to operate in batch to exhaustion as evidenced by decline in gas production.

Reactor feed was initiated at an organic loading rate (OLR) of 2.7 g COD·L⁻¹·d⁻¹ and a corresponding HRT of 11.5 days. The reactor was brought to the target operational HRT of 3.8 days in accordance with the schedule outlined in Table 3-2. The target HRT was selected on the basis that full scale IBRs are typically operated at HRTs of 3.5-4 days.
### Table 3-2: Reactor loading plan

<table>
<thead>
<tr>
<th>Time (days)</th>
<th>OLR (g COD/(L·d))</th>
<th>HRT (days)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-50</td>
<td>0</td>
<td>--</td>
</tr>
<tr>
<td>51-95</td>
<td>2.7</td>
<td>11.5</td>
</tr>
<tr>
<td>95-126</td>
<td>4.0</td>
<td>7.6</td>
</tr>
<tr>
<td>126-194</td>
<td>8.0</td>
<td>3.8</td>
</tr>
</tbody>
</table>

**Active Digestion Study Parameters:**
**Gas Production and Settleable Solids**

Significant energy input to the reactors is limited to heat input and biogas evolution. The most important anticipated differences between the clean water control studies and the active digestion studies were assumed to be the enhancing impact of evolved biogas bubbling out through the sludge bed and water column and the generation of biosolids in the reactor. Gas production was measured continuously for each active digestion treatment (Table 3-3). Reactor contents were evaluated to determine the settleable solids fraction per AWWA Imhoff Cone method (APHA-AWWA-WEF, 2005).

### Table 3-3: Energy input as biogas production and solids content for treatments. C = Clean water trial, L = lithium, Ra = Rhodamine WT trial I, Rb = Rhodamine WT trial 2 (washout), 3 = 35°C, 4 = 45°C, 5 = 55°C.

<table>
<thead>
<tr>
<th>Treatment</th>
<th>Biogas Production (e), L×d-1</th>
<th>Settleable Solids (mL/L)a</th>
<th>COD removal (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>C3</td>
<td>--</td>
<td>--</td>
<td>--</td>
</tr>
<tr>
<td>C4</td>
<td>--</td>
<td>--</td>
<td>--</td>
</tr>
<tr>
<td>C5</td>
<td>--</td>
<td>--</td>
<td>--</td>
</tr>
<tr>
<td>L3</td>
<td>222 (3)</td>
<td>500</td>
<td>56</td>
</tr>
<tr>
<td>L4</td>
<td>160 (23)</td>
<td>270</td>
<td>37</td>
</tr>
<tr>
<td>L5</td>
<td>251 (8)</td>
<td>120</td>
<td>59</td>
</tr>
<tr>
<td>Ra3</td>
<td>214 (12)</td>
<td>500</td>
<td>71</td>
</tr>
<tr>
<td>Ra4</td>
<td>184 (21)</td>
<td>270</td>
<td>41</td>
</tr>
<tr>
<td>Ra5</td>
<td>244 (10)</td>
<td>120</td>
<td>66</td>
</tr>
<tr>
<td>Rb3</td>
<td>222 (3)</td>
<td>500</td>
<td>56</td>
</tr>
<tr>
<td>Rb4</td>
<td>160 (23)</td>
<td>270</td>
<td>37</td>
</tr>
<tr>
<td>Rb5</td>
<td>251 (8)</td>
<td>120</td>
<td>59</td>
</tr>
</tbody>
</table>

* Settleable solids measured per AWPHA Imhoff cone method upon reactor decommissioning on 5/5/9

**Sample Measurement**

Fluorescent tracers were analyzed by pipetting well mixed samples into 96 well black opaque plates
(Corning Costar) and analyzing the samples in a BioTek (Winooski, VT S4MLFPT) plate reader fluorimeter with Gen 5 software (v1.04.5). The tungsten lamp was used with an excitation wavelength of 530 nm, and an emission wavelength of 585 nm. Sensitivity was set to 100, with a top probe vertical offset of 4.0 mm and a column offset of 0 mm. Lithium tracer samples were analyzed by the Utah Veterinary Diagnostic Laboratory (Utah State University, Logan, UT) using internal standard operating procedure 1245.0 for ICP/MS.

During the experiments, the reactor was continuously monitored for gas production. Grab samples for total solids (TS), volatile solids (VS), fixed solids (FS), and total COD were taken daily. Solids concentrations were measured in accordance with Standard Methods (APHA-AWWA-WEF, 2005). COD was measured by the closed reflux colorimetric method (ibid.). Biogas flow rates were recorded at 6 Hz using Labview 8.2 software (National Instruments, Austin, TX).

DATA REDUCTION

Data was reduced in accordance with the methods first proposed by Danckwerts (1953). The C vs T curve for each treatment was first plotted as a cumulative residence time distribution to determine the overall shape of the curve and develop a preliminary understanding of reactor hydrodynamics. Each C vs T curve was normalized to F(0) (fraction of maximum concentration) vs τ (fraction of HRT). This permits direct comparison of cumulative RTD (cRTD) curves for any reactor, including idealized curves from nonlinear curve fitting for interpretation of data.

These curves were examined visually to determine the most appropriate approach to further analysis. It is evident that the curves more closely approximate CSTR than PF behavior. The dispersion model, best used to represent systems with significant PF behavior, was eliminated as an approach to quantifying the reactor behavior, and the curves were analyzed using the tanks in series and combined model approaches (Fogler, 2006; Levenspiel, 1993, 1999b; Wen and Fan, 1975).
RESULTS AND DISCUSSION

RESIDENCE TIME DISTRIBUTION, CLEAN WATER STUDY

At the beginning of the experiments, dye was observed accumulating in the bottom ~10% of the reactor during the initial switch over to the dye feed; subsequent measurements of reactor temperature showed the existence of a thermocline at that level in the reactor that may be a function of the heating system in these particular reactor configurations.

F-τ and C-T curves are shown for all the three clean water studies in Figure 3-4, superimposed on both ideal plug flow and CSTR response curves. From the figure it is apparent that the reactor behavior more closely approximates CSTR behavior than plug flow. The curves show definite PF behavior at 0-10% HRT, but overall behavior is CSTR. While the RTD indicates the potential for plug flow in series with CSTR, PF can be eliminated as the primary fit model.

Figure 3-4: Cumulative RTD expressed as F-τ and C-T curves for clean water studies (no active digestion, Rhodamine WT tracer, 35°, 45°, and 55° C).

Fluctuations in individual treatments about the ideal curve are serially correlated as would be expected from time series data, indicating that they may due to real effects, not just measurement error which would be randomly distributed. Correlations between treatment dataset trends at 1.4<τ<1.7 and 2.0<τ<2.3 are puzzling and may be real effects or may be a result of insufficient randomization in the data analysis. They
do not however impact the overall conclusion that the reactors can be modeled as a PF and CSTR in series for the case of energy input being limited to heat flux through the reactor and the interaction of the resulting density and viscosity gradients.

From the graph it is evident that the fluid portion of the reactor behaves as a real complete mix system. A real complete mix system can be differentiated from the ideal complete mix system in terms of a two-parameter compartment model (Figure 3-5) that includes components of a stirred tank reactor, a bypassing flow that short circuits the reactor by traveling directly from the inlet to the outlet with minimal mixing, and a dead zone where no mixing occurs. Equation 2 represents the $F(\theta)$ function for the reactor model (Wen and Fan, 1975).

![Diagram](image)

**Figure 3-5: Real CSTR with bypass and dead space.**

The data was evaluated using the R statistical computing language (R Development Core Team, 2008) to evaluate Equation 2 using the nls package for nonlinear least squares analysis to solve for the parameters $v_1$ and $b$ (Table 3-4). Although difficulties with calibrating the Rhodamine tracer prevented the closing of the mass balance for the model and thus reliable application of $v_1$ and $b$ values to real systems, the data can be compared within the context of these experiments.

$$\frac{c_{e}}{c_{i}} = F(\theta)_{k} = 1 - \frac{v_1}{v} e^{-\left(\frac{\theta}{v} \frac{v_1}{b}\right)}$$

(2)
Where:
\( c_o \) = observed tracer concentration at time \( t \);
\( c_m \) = Maximum tracer concentration;
\( V_i \) = Volumetric flow rate through mixed volume;
\( V \) = Total volumetric flow rate;
\( \tau \) = Normalized time, \( \frac{t}{\theta} \);
\( b \) = Volume fraction of perfect mixing;
\( \theta \) = Hydraulic retention time (HRT), \( \frac{V}{V} \);
\( V \) = Total reactor volume;
\( d \) = Dead volume fraction = \( V \cdot b \)

<table>
<thead>
<tr>
<th>Treatment</th>
<th>R Label</th>
<th>( V_i )</th>
<th>b</th>
<th>RSS</th>
<th>CSTR ( \text{vol (%)} )</th>
<th>Plug ( \text{vol (%)} )</th>
<th>Dead ( \text{vol (%)} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Combined (all data)</td>
<td>cos</td>
<td>0.986</td>
<td>1.019</td>
<td>1.396</td>
<td>102</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Clean water 35(^\circ)C</td>
<td>c3</td>
<td>1.106</td>
<td>1.193</td>
<td>0.2713</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Clean water 45(^\circ)C</td>
<td>c4</td>
<td>1.0869</td>
<td>1.15</td>
<td>0.2391</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Clean water 55(^\circ)C</td>
<td>c5</td>
<td>1.1228</td>
<td>1.147</td>
<td>0.09307</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Clean water combined</td>
<td>coc</td>
<td>1.1051</td>
<td>1.163</td>
<td>0.6143</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Active lithium 35(^\circ)C</td>
<td>L3</td>
<td>0.9947</td>
<td>0.8041</td>
<td>0.3236</td>
<td>80</td>
<td>0</td>
<td>19</td>
</tr>
<tr>
<td>Active lithium 45(^\circ)C</td>
<td>L4</td>
<td>1.0243</td>
<td>0.6824</td>
<td>0.03292</td>
<td>68</td>
<td>2</td>
<td>30</td>
</tr>
<tr>
<td>Active lithium 55(^\circ)C</td>
<td>L5</td>
<td>0.9716</td>
<td>0.8496</td>
<td>0.02045</td>
<td>85</td>
<td>0</td>
<td>13</td>
</tr>
<tr>
<td>Lithium combined</td>
<td>coL</td>
<td>0.9948</td>
<td>0.7768</td>
<td>0.4127</td>
<td>78</td>
<td>0</td>
<td>22</td>
</tr>
<tr>
<td>Active RWTa 35(^\circ)C</td>
<td>ra3</td>
<td>1.028</td>
<td>0.9333</td>
<td>0.03769</td>
<td></td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>Active RWTa 45(^\circ)C</td>
<td>ra4</td>
<td>1.0509</td>
<td>1.245</td>
<td>0.06913</td>
<td></td>
<td>6</td>
<td></td>
</tr>
<tr>
<td>Active RWTa 55(^\circ)C</td>
<td>ra5</td>
<td>1.0228</td>
<td>1.195</td>
<td>0.07128</td>
<td></td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>RWTa combined</td>
<td>cora</td>
<td>1.031</td>
<td>1.122</td>
<td>0.28</td>
<td></td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>Active RWTb 35(^\circ)C</td>
<td>rb3</td>
<td>0.8997</td>
<td>1.1619</td>
<td>0.06274</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Active RWTb 45(^\circ)C</td>
<td>rb4</td>
<td>0.9384</td>
<td>1.1639</td>
<td>0.07251</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Active RWTb 55(^\circ)C</td>
<td>rb5</td>
<td>0.9689</td>
<td>1.173</td>
<td>0.04349</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>RWTb combined</td>
<td>corb</td>
<td>0.926</td>
<td>1.162</td>
<td>0.1839</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>35(^\circ)C combined</td>
<td>co4</td>
<td>1.0023</td>
<td>1.028</td>
<td>0.5442</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>45(^\circ)C combined</td>
<td>co5</td>
<td>0.9898</td>
<td>1.0696</td>
<td>0.2739</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>55(^\circ)C combined</td>
<td>co3</td>
<td>0.9741</td>
<td>0.9532</td>
<td>0.5402</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The plot of the NLS models generated in R for the clean water data are shown in Figure 3-6. The models show no important difference between treatments with approximately 10% of the reactor volume acting as plug flow, and the remainder as CSTR. When NLS curves are shifted to the origin to
accommodate the PF fraction, the remaining portion of the reactor volume tracks the ideal CSTR curve well.

![Graph showing F(θ) vs. t]  

Figure 3-6: Nonlinear least squares (NLS) models for clean water tracer studies including base model (L) and shift to origin to show fit with CSTR (R).

Quantile-Quantile plots of clean water tracer NLS residuals against the normal distribution (Figure 3-7) show good agreement with the normal distribution. Given this and the similarity of the NLS generated curves they are assumed to have come from the same distribution and the datasets are combined in a composite clean water model (Comp CW). From these data, it appears that when temperature flux is the only source of energy in the system, there is no discernable difference between mixing patterns in the IBR for the three temperatures studied.

![Graph showing Q-Q Normal plots]  

Figure 3-7: Q-Q Normal plots for clean water studies.
RESIDENCE TIME DISTRIBUTION: ACTIVE DIGESTION STUDIES

The clean water studies indicated the real CSTR model as previously discussed and helped confirm the approach used for the R analysis tools. PF was eliminated as a candidate for the primary mixing behavior in the IBR. This in turn eliminates the dispersion model as a valid predictor of reactor behavior (Levenspiel, 1993). Two active digestion studies were conducted with RWT and a third was conducted with lithium to provide a check on the RWT study results.

Assuming that the thermal mixing component is consistent regardless of reactor temperature, observed differences in mixing for treatments when grouped by tracer are due to the impacts of gas evolution/bubbling and settleable solids content. Given that in real systems, different combinations of gas production volume and solids content will be encountered, it is also reasonable to assume that all the active tracer data can be evaluated as if from a common distribution and compared as a function of temperature, tracer, or both. In all cases, it appears that gas induced mixing is sufficient to eliminate the plug flow volume in the reactor.

Lithium

F-t and C-T curves are shown for all the active Lithium studies in Figure 3-8, superimposed on an ideal CSTR response curve. Energy input from active digestion (gas evolution) appears to be sufficient to eliminate the PF volume identified in the clean water study. The cRTD curves show behavior as real-CSTR with potential for dead volume as evidenced by initial slopes higher than that of the ideal curve. Behavior is still CSTR, although the serially correlated deviations from smooth curves first observed in the clean water studies are also evident in this work and indicate that the system is subject to random events of short circuiting, recovery, and possibly sequestration that result in a distribution of possible concentrations with respect to time that average to CSTR behavior. For example, the L3 data shows an event (short circuiting) from 1.2<\(t<3<1.8\) that appears to be independent of the overall trend. From 2.5<\(t<3<1.8\), the system recovers and stabilizes at an equilibrium where \(C = C_0\). It is hypothesized that this may indicate an event like gas evolution and bubbling carrying a relatively large quanta of tracer directly through the hole in the septum, preventing incorporation into mixed volume and consequent loss of that mass of tracer to the
system. It was observed through the clear reactor walls that gas had a tendency to accumulate within the sludge bed until the buoyant force of the gas overcame the weight of the overlying sludge, causing a violent upwelling of gas and solids in the liquid volume that could conceivably produce vectors that would pass through the septum opening and upset the normal flow regime, skewing cRTD results. Ultimately, the upset would be rectified as tracer is diluted in the upper volume until it reaches the predicted equilibrium concentration.

Figure 3-8: Cumulative RTD expressed as F-τ and C-T curves for active digestion studies (lithium tracer, 35°, 45°, and 55° C).

The lithium data was processed in R to generate F vs τ curves for the NLS model (Figure 3-9). As a conserved tracer, the NLS data was used to quantify the dead space in the reactor (Table 3-4). L4 exhibited the most dead space, which is consistent with the amount of gas production in the reactor (Table 3-3); less gas would seem to yield less mixing. There appears to be no relationship between temperature/settleable solids/gas production (all confounded in study) and mixed volume given the visual correlation of the L3 and L5 curves with their similar gas production rates, but large disparity in solids content. Rb studies were conducted simultaneously, and theoretically should have shown similar behavior for the three treatments, but they did not. The dead volume implied by the NLS models may therefore be real, or it may be an artifact of the characteristics of the tracer and random bypass events that remove tracer
from the working volume without incorporating it into the system; the response curve then drops as concentrations in the upper chamber equalize with those in the larger chamber, implying dead volume and bypass when they are events that occur in the reactor, but are not consistent characteristics of the reactor.

![Graph](image)

Figure 3-9: NLS model output for lithium tracer under active digestion conditions. L4 shows more dead space (30%) than L3 or L5, as might be anticipated given the lower level of energy input from gas mixing (Table 3). The differences between L3 and L5 cannot be considered to be important given the confounding of the gas production and solids content factors.

**Rhodamine WT, First Active Digestion Study (RWTa)**

F-τ and C-T curves are shown for the first group of active digestion Rhodamine WT studies in Figure 3-10, superimposed on an ideal CSTR response curve. RTD curves show the possibility of tracer sorption or reaction with the Ra4 and Ra5 treatments, while Ra3 tracks the ideal CSTR curve closely. Ra3 has the most accumulated biomass, and thus the most potential for interaction which should manifest as tracer loss.

The Ra data was processed in R to generate F vs t curves for the NLS model (Figure 3-11). Ra3 treatment tracks the ideal very closely while both Ra4 and Ra5 appear to show some tracer loss. As with the Li and CW studies, there appears to be no consistent pattern of behavior between treatments as a function of temperature. The R3 treatment implies some dead volume, consistent with the lithium treatments, but the others do not. The dead zones may therefore be real for all system components, or they may be specific to the Lithium tracer and a function of interaction with the reactor fluid.
Figure 3-10: Cumulative RTD expressed as F-τ and C-T curves for first set of Rhodamine active digestion studies at 35°, 45°, and 55° C.

Figure 3-11: NLS model output for first run (Ra) Rhodamine WT tracer under active digestion conditions.

Rhodamine WT, Second Active Digestion Study (RWTb)

F-τ and C-T curves are shown for the second group of active digestion Rhodamine WT studies superimposed on an ideal CSTR response curve in Figure 3-12. Study results were from a washout (elutriation) study measuring decline in Rhodamine WT concentration simultaneously with the Lithium RTD study. Monitoring the decline in tracer concentration as it is replaced by fresh water produces an I
curve, which can be converted to an F curve by the methods outlined in Danckwerts (1953). Assuming tracer interactions are similar, RTDs should have also been comparable between Rb and L treatments. They were not. While there is some slight indication of dead space in the F3 curve, the rest of the curve is below the ideal, demonstrating a potentially nonconservative tracer. The F5 curve is inconclusive for $\tau < 1$ given contaminated samples in this range. Overall, the samples appear to follow the CSTR curve well, although they do show evidence of tracer loss.

![Figure 3-12: Cumulative RTD expressed as F-τ and C-T curves for second set of Rhodamine active digestion studies at 35°, 45°, and 55° C. Studies observed decline in tracer concentration as a washout study. F and C curves are developed from I curve.](image)

The Rb data was processed in R to generate F vs $\tau$ curves for the NLS model (Figure 3-13). NLS models with y-intercepts greater than 0 generally indicate bypass as a reactor characteristic, but given that none of the other treatments showed this behavior, this is regarded as unlikely. The curves do not reveal obvious differences in the models. Temperature effects do not appear to have an effect on mixing.
Treatments as a function of Temperature

When the studies were compared as a function of temperature (Figure 3-14), Lithium consistently showed higher levels of dead space in reactors than Rhodamine. Rhodamine studies are largely indistinguishable at 45 and 55°C. Overall, the randomness and lack of correlation between minor events in treatments is evident over the course of the studies. There is clear serial correlation within treatments, but beyond this each treatment produces distinctly different patterns in the response curve, whether grouped by temperature or tracer type.

As with the scatter plot data, the NLS models (Figure 3-15) clearly show the Li tracer’s tendency to predict dead space when correlated by temperature, while the Rhodamine studies tend to demonstrate loss of tracer. The composite of the three treatments for each temperature levels, generated by using NLS to fit the data for all three treatments to a single model shows very good agreement to the ideal CSTR at all temperature levels.
Figure 3-14: Cumulative RTD expressed as F-τ curves for all active digestion treatments grouped by temperature.
Combined Studies

Figure 3-16 shows all the data points for all active digestion treatments superimposed on a plot of the idealized CSTR curve. The data implies reasonably good fit to the ideal CSTR, but with significant potential for deviation from the ideal.

Combining all the data into one set and fitting the data to the model in NLS yields the curve shown in Figure 3-17. The figure also shows the 95% confidence interval that bounds the model. The generated model is in very good agreement with the ideal CSTR model. It is recognized that there are clear
differences between tracer treatments; there is dead volume with respect to the lithium tracer, but it is not clear as to why this effect was not observed for the Rhodamine tracer. Possible explanations include the aforementioned difficulty in establishing calibration for the RWT or tracer loss from the system. With the relatively balanced fit around the ideal for all individual treatments, it appears to be reasonable to assume that the IBR produces a CSTR behavior with important variability in mixing quality.

Figure 3-16: Combined data for all treatments plotted against ideal CSTR.

Figure 3-17: NLS F-curve generated from composite of data from all treatments. Curve is nearly indistinguishable from ideal CSTR. 95% confidence interval bounds shown with overlay of data used to generate model.
Another check on the validity of combining the data can be made by plotting the residuals of the NLS generated model against a normal distribution as quantile-quantile plots to observe the normality of the residual distribution. As Figure 3-18 shows, the data compares well with the normal distribution. There are clear outliers, but most of the residuals approach a normal distribution. Notable exceptions include the 45°C composite model residual and the Rhodamine A composite model, both of which show some tailing, but the majority of the data is linear with the 45°C line on the plots. The overall composite model also exhibits tailing, but again, the majority of the data points fall on the 45°C line, indicating good agreement with the normal distribution and enforcing the assumption of normality. By combining the data, it is assumed that the random effects of individual RTD curves are balanced, and the effects characteristic of tracer interactions with the system are confounded with the net result being a large dataset that more accurately reflects the average response of a complex system.

Figure 3-18: Matrix of Q-Q Normal plots for all active digestion NLS models.
TANKS IN SERIES MODEL

Reactors were also evaluated using the CSTR in series model in accordance with the methods outlined in Fogler (2006) and Tchobanoglous et al. (2003), and as summarized in Table 3-5. Any non-ideal reactor can be represented as a set of ideal CSTRs operating in series. The number of CSTRs in series is equal to the inverse of the dimensionless variance of the system, \( \sigma^2 \). \( \sigma^2 \) is equal to the variance of the C-T curve divided by the theoretical hydraulic retention time. For a comprehensive discussion of the techniques involved, see Section 14.2 in Fogler (2006).

Table 3-5: Summary of Calculated Parameters for CSTR in Series Model

<table>
<thead>
<tr>
<th></th>
<th>CW</th>
<th>Li</th>
<th>RWTa</th>
<th>RWTb</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \theta ) (HRT), d</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>C3</td>
<td>C4</td>
<td>C5 Avg</td>
<td>L3</td>
</tr>
<tr>
<td>t_{ave} (RTD HRT), d</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CSTR in series</td>
<td>1.44</td>
<td>1.60</td>
<td>0.89</td>
<td>1.31</td>
</tr>
</tbody>
</table>

All the trials produced results indicating that the IBR behaved as 1 to 2 CSTRs in series. This is consistent with the results of the multi-compartment model wherein it was shown that the behavior approximates a real CSTR with some inefficiencies. It is acceptable to use non-integer values for representation of the number of tanks in series; the average number of CSTRs required to represent the IBR in these trials is 1.2. Solving for conversion efficiency using this approach is a matter of solving for conversion at 1 CSTR and at 2 CSTRs, thus providing an upper and lower limit of conversion.

CONCLUSIONS

Lithium and Rhodamine WT were used to investigate the mixing behavior of the Induced Bed Reactor (IBR) anaerobic digester. Step input studies in both clean water and active digesters showed that liquid fraction of the IBR operates as a complete mix reactor with potential for deviation from ideal CSTR behavior when analyzed using both the combined and CSTR in series models.

Energy input in the clean water tracer study was limited to the heat required to maintain temperature in
an insulated reactor. This represented the minimum energy input to the system and demonstrated that the heat flux creates sufficient fluid movement in the reactor to mix the contents over the HRT studied with a plug flow component equal to approximately 10% of the reactor volume. The clean water study shows no important difference in mixing as a function of reactor temperature.

Based on the results of the clean water RTD studies, the dispersion model was deemed inappropriate to represent the behavior of the IBR given its proximity to CSTR behavior, and a real CSTR compartment model with elements of bypass and dead space was selected for further investigation. This model was applied to the tracer data along with the CSTR in series model. The conclusion of the CSTR in series model is similar in that it considers the IBR to operate as 1.2 CSTRs in series.

The RWT fluorescence was difficult to calibrate to concentration in the active digestion fluid. A simple self referencing 2-point calibration was used, assuming that the maximum detected fluorescence was representative of the maximum concentration. This permits qualitative analysis of the normalized cRTD curves, but does not permit quantitative determination of compartment sizes in the multi-compartment model. The authors recommend that future researchers considering Rhodamine WT for RTD studies in an active anaerobic digester proceed with caution if mass balance closure is required. Conversely, the lithium tracer was relatively simple to calibrate, and was deemed reliable for use in generating a mass balance which could then be used to evaluate the size of dead zones and bypass flows for a given dataset. Since some of the RTD characteristics predicted by the lithium curves were not reflected in the Rhodamine curves, it appears that there may be differences in the way that the tracers reflect mixing behavior in the IBR content matrix.

The lithium study implies a dead volume at all temperature levels, with dead volume being correlated with gas production as might be expected. This behavior was not corroborated by the Rhodamine studies, and is regarded as possible but inconclusive. The Rhodamine studies indicate CSTR behavior. The Rhodamine curves appear to show that tracer may be consumed in the system, but this may also be a function of the calibration method.

All the studies clearly show that while approaching the ideal of a CSTR on average, mixing in the IBR is subject to events that may impact effluent quality by providing varying degrees of mixing at apparently
random intervals. The closure of the RWT mass balance would have been beneficial, but the qualitative results showing that while the reactors are not ideal, they can be viewed on average as CSTRs and still provide significant insights into reactor behavior. There are irregularities in the flow and mixing patterns even under the laboratory conditions used in this work that will probably be exacerbated in real world applications where reactors will be subjected to more complex substrates and less environmental control. These irregularities are thought to be inherent to the low-energy input reactor design. While the biological mass balance model of the IBR can consider the reactor as a CSTR, in order to minimize the impact of the uneven mixing inherent to the design in practice, multiple reactors in parallel or series should be considered to help to normalize the mixing distribution and thus the reactor performance and effluent quality.

Finally, while this study addresses the hydrodynamic mixing behavior of the IBR as being a non-ideal CSTR, the IBR is a retained biomass reactor, and the assumption that HRT = SRT that generally follows CSTR designation does not apply. SRT must be calculated independently using established methods.

REFERENCES


CHAPTER 4

PERFORMANCE OF THE INDUCED BED REACTOR ANAEROBIC DIGESTER AT
MESOPHILIC AND THERMOPHILIC TEMPERATURES

The Induced Bed Reactor (IBR) was developed at Utah State University (USU) to apply high-rate anaerobic digestion techniques to high-suspended solids content substrates (3-12% total solids) such as food waste and dairy manures. This technology has been successfully implemented at full-scale multiple installations in the United States and Canada as a waste treatment and energy production technology.

58L bench scale reactors were operated at 35° C, 45° C and 55° C under three organic loading rates and three corresponding hydraulic retention times for each reactor using a dairy manure starter culture and a dextrose/yeast extract substrate at 30.6 g/L COD. Influent and effluent streams were monitored for parameters including solids composition, VFAs, gas quality and quantity, and chemical oxygen demand (COD). Results were compared with a previously published study on operation of the IBR at thermophilic temperatures (55° C).

The IBRs were successfully operated for over 180 days, demonstrating a peak COD removal rates of 89% at 35° C. Development of granulated sludge beds comprising settled sludge volumes of 500 mL-L⁻¹, and 250 mL-L⁻¹ was evident. The IBR was demonstrated to operate at all three temperature levels, although the 45° C reactor was susceptible to process upset, and the 55° C reactor produced consistently poor quality gas compared to the other reactors.

Gas samples collected and stored in Tedlar bags proved to be consistently and readily contaminated, apparently by diffusion of O₂, CO₂ and N₂ through the bag wall, pointing to the necessity of developing a standard for collection and analysis of biogas if Tedlar bags are to be used for capture and storage.

INTRODUCTION

The purpose of anaerobic digestion in wastewater treatment is to reduce waste sludge volume and activity by fermenting the waste in the absence of oxygen. Anaerobic digestion relies on the symbiotic

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¹ Coauthored by J. S. Dustin and C. L. Hansen
relationship between two general classes of anaerobic microorganisms to catabolize carbonaceous substrates to the relatively stable and innocuous end products of biosolids and biogas (Bryant et al., 1967).

The relatively long doubling times of anaerobes require either long residence times (20-40 d) to allow populations to accumulate to levels permitting significant treatment or separation of hydraulic retention time (HRT) and solids residence time (SRT) to permit high-rate treatment (Tchobanoglous et al., 2003). Methods for accomplishing this include separating the active solids from the effluent and recycling them to the digester, by utilizing an attached growth process, or through the use of an induced bed or blanket of active biosolids that self-select to remain in the reactor as the substrate passes through (Gerardi, 2003; McCarty, 1981). These high rate processes generally require a relatively low suspended solids concentration in the influent to avoid dilution of the active biomass and to avoid operational problems in the digester (plugging, sedimentation, short circuiting, etc.) (Tchobanoglous et al., 2003).

The inability to treat wastes with high influent suspended solids content poses a problem for application of high rate processes to substrates like those associated with food processing wastes and livestock manures (Chen, 1986; Schofield, 1984). Processes addressing these wastes therefore commonly rely on plug flow or CSTR configurations with no solids retention or recycle as a reactor designed to retain or recycle solids will necessarily accumulate influent solids as a function of design.

In the early 1990s, Dr. Conly Hansen at Utah State University proposed a digester configuration, the Induced Bed Reactor (IBR) (Hansen and Hansen, 2005), that was designed to address the operational limitations of applying high rate treatment techniques to high strength, high solids agricultural substrates. The IBR relies on gravimetric biomass retention in the digester to provide an accumulation of active solids, similar to the operational principles first proposed by Lettinga et al. (1980) for the UASB reactor design (Figure 4-1). UASBs operate on the principle of separation of SRT and HRT via granulated biomass retention. The active biomass acts as an effective filter, metabolizing available carbon to CO₂ and CH₄. In the UASB, influent enters the reactor from the bottom via a diffuser to provide equal distribution through the sludge bed and minimize short-circuiting. Substrate transits the sludge bed vertically, converting chemical oxygen demand (COD) to biogas. Biogas is directed via the deflector baffles to the 3-phase
separator, while treated liquid effluent flows around the baffles and the separator and over a weir to the reactor outlet.

Figure 4-1: Cutaway view of an Upflow Anaerobic Sludge Blanket (UASB) reactor (L) and an Induced Bed Reactor, IBR (R).

In 1993, Dr. Hansen began working with digestion of high-suspended solids, and high strength wastewaters in reactors configured as UASBs in an attempt to translate some of the advantages of the high-rate small footprint systems to agricultural and food wastes. This work resulted in the development of the Induced Bed Reactor (IBR) (Figure 4-1), an upflow design that preserves some of the biomass concentrating characteristics of the UASB while allowing for the processing of complex high solids and wastewaters such as dairy manures by also permitting the passage of solids through the reactor. Key differences between the UASB and the IBR include the septum baffle which separates the mixed liquid volume from the headspace, the diffuser plate which provides substrate diffusion into the sludge bed, and the auger which keeps the gas outlet in the septum baffle clear of solids. Liquid effluent exits the reactor via an inverted trap above the septum.
These differences are derived from the need to address the high solids content of the influent. Inlet and outlet conditions and the gas separation mechanism were modified to minimize plugging in the reactor. The principal advantages of the IBR system are that the system a) reduces the need for solids separation over what would be required for other high-rate systems, b) with a 4-day hydraulic retention time (HRT) requires 5x less volume than a similarly performing plug or complete mix digester with a 20-d HRT, and c) permits the thermal processing of suspended solids in the influent at thermophilic temperatures. At thermophilic temperatures, though recalcitrant suspended solids in the influent may not be subject to digestion, they can pass through the reactor where they are raised to thermophilic temperatures and gain the benefit of pathogen destruction before being carried out with the effluent stream.

In defining contacting patterns and mixing behavior in the IBR, previous research by the authors has shown that the IBR can be treated as a completely stirred tank reactor (CSTR) with elements of dead space. The reactors in series model indicates that the reactor approximates 1.2 CSTRs.

The IBR has been successfully implemented on a commercial scale with over 4,000 m$^3$ of installed capacity in the US and Canada, but there is very little published data on operation of the system. The purposes of this study were to observe the IBR at a range of temperatures and loading rates and gather data on operational behaviors, conversion performance, and biomass characteristics to permit development of a mass balance model that describes the behavior of the digester under the conditions studied. The key operating parameters examined were solids production, methane production, and COD reduction. The study focused on reactor behavior at an HRT of 3.8 days given that most full-scale IBRs operate at HRTs of 3.5-4 days.

**MATERIALS AND METHODS**

**EXPERIMENTAL APPARATUS**

The experimental apparatus for the research is shown in Figure 4-2. Bioreactors with 58-L wetted volume were constructed from extruded acrylic tubing 0.30 mm ID and 0.91 m high. Influent feed was controlled by peristaltic pumps (Masterflex 7523, Cole-Parmer, Inc., Vernon Hills, IL). During the experiments, the systems were maintained at 35° and 45° C by heat tape applied to the bottom 0.3 m of the reactor and controlled by digital temperature switches (Love TS-13011, Dwyer Instruments, Michigan City,
IN) utilizing thermistors placed in thermowells at the mid-height of the reactors. The reactors were wrapped with 0.1 m of fiberglass insulation. Gas production was measured via thermal mass flowmeters (822S-L-2-ON1-PV1-V4, Sierra Instruments, Monterey, CA).

![Diagram of the research apparatus](image)

**Figure 4-2: Schematic design of the research apparatus. Three identical systems were constructed and operated simultaneously for the duration of the experiments.**

Substrate was refreshed in the feed tank at 7-10 day intervals, and maintained in a mixed state using a variable speed stand mixer operated at 1800 RPM (EV1P25M1C48, Lightnin, Rochester, NY). The feed tank was maintained in a cold room at 5°C to preserve substrate quality. For each reactor, the feed pump was set to provide a constant rate feed to the digester at the desired HRT. Foam control was provided by semi-continuous injection of Sigma Antifoam B (Sigma Aldrich, Inc. St Louis, MO) via diaphragm metering pumps (LMI AA97, Milton Roy Americas, Ivyland, PA) to maintain an antifoam concentration in the upper chamber of the reactor of 30 mg/L. Effluent gas was routed through liquid capture foam traps, thence to refrigerated water columns where excess water vapor was removed by bubbling the gas through the column at 5°C before warming the gas by circulating it through tubing at room temperature to 22°C for flow measurement. Excess gas was vented via a fume hood.
FEED SUBSTRATE AND INOCULATION

The substrate solution had a chemical oxygen demand (COD) of 30,600 mg/L and was composed of 23,438 mg/L of dextrose, plus the following nutrients (mg·L⁻¹): 2,500 yeast extract, 2,656 NH₄Cl, 525 K₂HPO₄, 225 FeCl₂ · 4H₂O, 469 CaCl₂ · 2H₂O, 391 MgSO₄ · 7H₂O, and 313 KCl. To prevent required microbial trace element deficiency, a trace nutrient solution (mg·L⁻¹) (500 H₃BO₃, 500 ZnCl₂, 300 CuCl₂, 5000 MnSO₄ · H₂O, 500 (NH₄)₆Mo₇O₂₄ · 4H₂O, 500 AlCl₃, 500 CoCl₂ · 6H₂O, and 500 NiCl₂) was added by 0.01% (v/v) to each batch of substrate. NaHCO₃, 10,000 (mg·L⁻¹), was added to maintain initial buffering capacity, and tap water (City of Logan, UT) was used as dilution water. The components were similar to those used for cultivating anaerobic bacteria in an anaerobic sequencing batch reactor (ASBR) by Cheong et al. (2007). Analysis showed that substrate influent pH was 8.0, while the remaining properties were maintained at the following values (g·L⁻¹): COD = 30.6, pH = 8.0, total solids (TS) = 22.2, volatile solids (VS) = 12.9, and fixed solids (FS) = 9.30.

Seed sludge taken from an operating IBR (Wade Dairy, Ogden, UT) was used to inoculate the reactors. The seed sludge was passed through an 8 mm mesh screen to remove larger solids, and the reactors were filled with a mixture of 1/3 sludge, 1/3 water, and 1/3 feed substrate. They were then brought to temperature over 5 days, and allowed to operate in batch until the bioavailable energy value of the substrate was exhausted as measured by the stabilization of biogas production.

LOADING RATE

The objective of the study was to maximize the time spent at the target HRT and OLR to enable extended observation of reactor performance and to support multiple tracer studies to permit characterization of mixing in the IBR. Changes in HRT were made when reactors had been running for approximately 4 HRTs at a given loading rate or when pH and gas production rates demonstrated relative stability for a minimum of 2 HRTs. Given that full scale IBRs are typically operated at HRTs of 3.5-4 days, the target HRT for this study was set at 3.8 days with a corresponding organic loading rate (OLR) of 8.0 g COD·L⁻¹·d⁻¹. The target loading rate was reached by maintaining a steady influent COD concentration of 30.6 g COD·L⁻¹ while decreasing the HRT in steps as outlined in Table 4-1. Table 4-2 shows the
controlled operating parameters for the experiments including the days that the reactor was operated in a steady state and the number of days and HRTs at steady state for each loading rate.

Table 4-1: Reactor loading summary

<table>
<thead>
<tr>
<th>Time (days)</th>
<th>OLR (g COD·L⁻¹·d⁻¹)</th>
<th>HRT (days)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-50</td>
<td>0</td>
<td>--</td>
</tr>
<tr>
<td>51-95</td>
<td>2.7</td>
<td>11.5</td>
</tr>
<tr>
<td>95-126</td>
<td>4.0</td>
<td>7.6</td>
</tr>
<tr>
<td>126-194</td>
<td>8.0</td>
<td>3.8</td>
</tr>
</tbody>
</table>

Table 4-2: Periods of operation and independent variables at steady state

<table>
<thead>
<tr>
<th>HRT (d)</th>
<th>35C</th>
<th>45C</th>
<th>55C</th>
</tr>
</thead>
<tbody>
<tr>
<td>HRT1</td>
<td>11.5</td>
<td>11.5</td>
<td>11.5</td>
</tr>
<tr>
<td>HRT2</td>
<td>7.6</td>
<td>7.6</td>
<td>7.6</td>
</tr>
<tr>
<td>HRT3</td>
<td>3.8</td>
<td>3.8</td>
<td>3.8</td>
</tr>
<tr>
<td>Steady State Operation (days)</td>
<td>58-91</td>
<td>68-91</td>
<td>58-89</td>
</tr>
<tr>
<td>Elapsed time @ Steady State (days)</td>
<td>33</td>
<td>23</td>
<td>31</td>
</tr>
<tr>
<td>HRTs @ Steady State</td>
<td>2.9</td>
<td>2.0</td>
<td>2.7</td>
</tr>
<tr>
<td>Influent OLR (g COD·L⁻¹·d⁻¹)</td>
<td>2.66</td>
<td>2.66</td>
<td>2.66</td>
</tr>
</tbody>
</table>

**Sampling**

Liquid influent samples were collected in the substrate feed tank. Effluent samples were collected 3-5 times a week at the reactor outlets. Gas samples were collected in 1-L Tedlar sample bags with polypropylene fittings (CEL Scientific, Santa Fe Springs, CA) for composition analysis. Composite samples for analysis of reactor contents at shutdown (in-situ suspended solids) were collected as grab samples by first gently agitating the reactor contents with a 50 mm x 1.5 m polycarbonate paddle to suspend solids, then taking multiple full column samples with a sludge sampler (Sludge Judge, Nasco, Fort Atkinson, WI) and recombining those as composite 1-L samples representative of each reactor.

**Analytical Procedures**

Parameters generally used in monitoring digester health and performance include hydraulic and organic loading rates, biogas and methane production rate, oxygen demand reduction, pH, and volatile fatty acid (VFA) concentrations (Wilke and Colleran, 1988). Effluent grab samples were taken 3-5 times a week and analyzed for COD, TS, VS, and FS in accordance with Standard Methods (APHA-AWWA-WEF, 2005).
COD was measured by the closed reflux colorimetric method. Grab samples were also taken from the midpoint of the reactors during the initial phases of the experiment until sample ports failed.

**Biogas**

Biogas flow rates were recorded at 6 Hz using LabView 8.2 software (National Instruments, Austin, TX). The methane content of the biogas was calculated from a COD mass balance at steady state, and checked against analysis by gas chromatography (GC, HP 6890 series, Hewlett-Packard, Wilmington, DE) using a capillary column (RT-Msieve 5A PLOT, Restek, Bellefonte, PA) with dimensions of 30.0 m x 320 μm x 30.0 μm. The column temperature was 35°C, while the inlet port and thermal conductivity detector temperatures were 50 and 200°C, respectively. Argon was used as the carrier gas at a flow rate of 1.5 mL/min. Gas standards were obtained from Scott Specialty Gases (Plumsteadville, PA) for calibration. COD consumption for methane production was calculated as 0.378 L CH₄/g - COD when measured at 22°C.

**Volatile Fatty Acids**

VFAs including acetate, propionate, butyrate, and valerate were measured via gas chromatograph (GC, HP 6890 series, Hewlett-Packard, Wilmington, DE), using a cross-linked polyethylene glycol capillary column (HP-INNOWax) with dimensions of 30.0 m x 250 μm x 0.25 μm. The temperature program for the column was 70 °C held for 15 min, 115 °C held for 3 min, and 240 °C held for 1 min. The inlet port and flame ionization detector temperature was 250 °C. Argon was used as the carrier gas at a flow rate of 1.8 ml/min.

**Suspended Solids**

In order to determine the critical parameter of solids retention time (SRT), active biomass was estimated as TSS. SRT was determined by constructing a mass balance around the reactor (Figure 4-3) using mass flow rates and calculating an SRT for each reactor at the ultimate loading rate based on measurements of in-situ and effluent TSS as shown in Equation 1 where $X_{INF}$ and $X_{INF}$ are mass flow rates, and $dX_{RET}$ is the total in-situ TSS in the reactor volume. With the completely dissolved, completely bioavailable substrate,
any suspended solids in the reactor or the effluent were assumed to be active biomass and \( X_{\text{inf}} \) was set equal to zero given that all influent solids for these experiments were dissolved by design.

![Diagram of Induced Bed Reactor](image-url)

**Figure 4-3: Schematic for mass balance model of Induced Bed Reactor. COD = chemical oxygen demand; \( X \) = suspended solids concentration**

\[
\text{outflow - inflow = retention} \\
X_{\text{EFF}} - X_{\text{INF}} = \frac{dX_{\text{RET}}}{dt} \\
\therefore \frac{dX_{\text{RET}}}{dt} = \frac{X_{\text{EFF}} - X_{\text{INF}}}{SRT}
\]

Reactor sludge beds were sampled upon completion of experiments by agitating the reactor contents and taking full column samples as previously described. The samples thus retrieved were analyzed for settleable solids volume using Imhoff cones in accordance with Standard Methods (APHA-AWWA-WEF, 2005). They were also compared qualitatively by rinsing 50 mL aliquots in petri dishes with distilled water to remove colloidal solids, then visually observing the sludge overlaid on a 2 mm grid.

**COD Mass Balance and Methane Content**

A mass balance based on the schematic presented in Figure 4-3 was constructed on COD at each pseudo steady state loading rate for each reactor (Equation 2). Criteria for defining steady state operation were 1) consistent biogas production rate, 2) consistent pH and 3) consistent effluent quality. Given that there is no prior data on IBR performance parameters, steady state periods were determined based on the visual correlation of observed output from reactor monitoring. Cyclical gas production rates were deemed to be
not significant in defining steady state behavior given that cycles can be correlated directly to the substrate replacement schedule and were considered to be a characteristic of biomass response to fresh vs. mature substrate for the study conditions.

\[
\text{inflow - outflow} = \text{retention} \\
\text{COD}_{\text{INF}} - (\text{COD}_{\text{LE}} + \text{COD}_{\text{BG}}) = \frac{d\text{COD}_{\text{RET}}}{dt}
\] (2)

In addition to colorimetric measurement as described previously, COD can be calculated based on the stoichiometry of the reactions required to completely oxidize reactor contents, and therefore provides a reliable common denominator for use in quantifying mass transfer through the system. COD has some of the same difficulties as TSS, namely that it must be carefully fractionated, and the more complex the substrate and the more accuracy is required in the results, the more care must be taken in determining the bioavailability of the COD in the system. Speece (1996) and Tchobanoglous et al. (2003) provide useful methodologies for making these determinations. As with the TSS, it was assumed that since all the COD entering the system was theoretically bioavailable and the reactor is essentially a CSTR, any COD retained in the system was retained as settleable solids (biomass), with the remainder being passed as dissolved COD in the effluent or converted to biogas and removed from the system.

Influent substrate (\text{COD}_{\text{INF}}) entered the system as a completely dissolved and directly measured fortified sugar mixture. The influent substrate was assumed to be completely bioavailable. Outflow, consisting of liquid effluent (\text{COD}_{\text{LE}}) and biogas (\text{COD}_{\text{BG}}), was also monitored regularly as outlined above. \text{COD}_{\text{LE}} was measured directly. \text{COD}_{\text{BG}} was calculated given that rate of accumulation in the reactor at steady state was, by definition, equal to zero (d\text{COD}_{\text{RET}}/dt = 0). The mass balance could then be solved for \text{COD}_{\text{BG}} using Equation 2 at the steady state periods outlined in Table 4-2. Biogas quality (bq), the unitless ratio of volume of \text{CH}_4 to volume of biogas was then calculated as outlined in Equation 3. Specific biogas production rate \( = \text{sbpr} = \text{L biogas produced} \cdot \text{L}^{-1} \cdot \text{wetted volume of reactor} \cdot \text{d}^{-1}.\)
\[ bq = \text{shpr} \times \frac{0.378 \text{L CH}_4}{\text{g COD}} \]

At non-steady state conditions, COD_{RET} can be related stoichiometrically to cell mass retention using the approximate formula for cell mass first described by Hoover and Porges (1952), \( \text{C}_5\text{H}_7\text{NO}_2 \) as outlined in Equation 4. The system mass balance can therefore also be used to determine the quantity of cell mass retained in the reactor as a function of retained oxygen demand.

\[
\text{C}_5\text{H}_7\text{NO}_2 + 5\text{O}_2 \rightarrow 5\text{CO}_2 + \text{NH}_3 + 2\text{H}_2\text{O}
\]

\[
\therefore \text{COD}_{cells} = \frac{\Delta(O_2)}{\Delta(C_5H_7NO_2)} = \frac{5 \times 32}{113} \times \frac{\text{g mol}}{\text{mol}} = 1.42 \times \frac{\text{g(O}_2\text{)}}{\text{g(cells)}}
\]

RESULTS AND DISCUSSION

ACTIVE GRANULATED SOLIDS AND SOLIDS RETENTION TIME

The reactors were operated for a total of 194 days in a laboratory setting. The insulation covering the reactors was periodically removed to permit visual observation of the biomass in the reactors and of the behavior and location of the sludge bed. At day 112, a crack in the 55° C reactor wall necessitated the replacement of the bottom half of the reactor with an opaque stainless steel patch, limiting visual observation of that reactor after that point. All three reactors demonstrated similar behavior, gradually developing sludge beds with a granular character. The beds were stratified, with larger sludge granules accumulating at the bottom of the bed and smaller granules and flocculated sludge towards the top of the bed. Beds rested on the bottoms of the reactors. The granules were interspersed with thin rod-shaped structures that had the appearance of clipped hairs 1-3 mm in length. Bed formation demonstrated that the reactor feed system utilized for the experiments was appropriate for developing a dense granular sludge.

Gas pockets could be observed forming in the sludge bed through the clear polycarbonate reactor walls, although a well attached black biofilm developed in the 55° C reactor, eventually rendering the wall opaque and restricting visual observation of reactor contents. Much of the observed biogas generated in the sludge
bed would make its way into the liquid volume of the reactor in the form of small bubbles less than 1 mm in size with minimal disruption of the sludge matrix.

Gas would also tend to accumulate in pockets visible through the reactor wall. Pockets as large as 6 cm long and 1 cm high were observed in the bed. When the buoyancy of a pocket overcame the force of the sludge on top of it, it appeared to float the biomass for a short distance until biomass cohesion was disrupted and the bubble would rapidly and violently ascend to the surface. The overlying biomass was dispersed into the liquid volume of the reactor with some of the denser material descending back into the displaced volume and the balance of the cavity being filled by adjacent material settling into the void space. The disturbance from these large eruptions of biogas could be observed impacting the full visible portion of the reactor in some instances. Granules and flocs served as markers to indicate bubble induced circulation patterns throughout the liquid column. The regularity of and apparent energy release from these events would appear to be a circumstantial corroboration of the results reported for the residence time distribution studies conducted concurrently with this work.

Flocculent sludge was also observed to accumulate in the upper chamber of the reactors above the septum. Spikes in effluent suspended solids and COD would occur periodically when these flocculated sludge beds would apparently reach a critical volume and begin mixing with the effluent. Impact of this upper bed on improvement of average effluent quality was not quantified, but for maintenance of consistent effluent quality, mixing of the upper chamber could be considered to homogenize the contents and eliminate the occasional spikes in effluent TSS and COD.

At shutdown, representative sludge samples were recovered and analyzed as previously described. Results are presented in Table 4-3. Settled sludge volumes as determined using Imhoff cones (SSV) were consistent with observations made through reactor walls while the reactors were in service. SSV for the 35°C reactor corresponded to a sludge bed occupying approximately half of the reactor volume; for the 45°C reactor, the sludge bed occupied approximately 25% of the reactor volume. The 55°C sludge bed could not be observed directly given the accumulated biofilm on the reactor walls and that the lower half of the reactor wall was covered with a stainless steel sleeve to repair a crack, but the measured SSV corresponded to approximately 12% of the reactor volume. These observations are also consistent with the
predictions of the van't Hoff-Arrhenius relationship which predicts an approximate doubling of reaction rate (and thus biogas production) for every 10°C increase in temperature. Biogas flow rates were roughly equivalent at each organic loading rate despite the differences in temperature, confirming that using in-situ TSS and biogas generation as indicators, the biomass activity in the systems was as would be expected.

Table 4-3: Sludge Characteristics, biogas flow rates, and Solids Retention Time (SRT). Standard deviations provided for mean values in parentheses.

<table>
<thead>
<tr>
<th></th>
<th>35°C</th>
<th>45°C</th>
<th>55°C</th>
</tr>
</thead>
<tbody>
<tr>
<td>HRT (d)</td>
<td>11.5</td>
<td>7.6</td>
<td>3.8</td>
</tr>
<tr>
<td>Settled Sludge Volume (mL·L⁻¹)</td>
<td>--</td>
<td>500</td>
<td>--</td>
</tr>
<tr>
<td>Effluent VSS (g·L⁻¹)</td>
<td>1.33</td>
<td>0.72</td>
<td>0.64</td>
</tr>
<tr>
<td>Final in-situ VSS (g·L⁻¹)</td>
<td>--</td>
<td>39.5 (6.31)</td>
<td>--</td>
</tr>
<tr>
<td>Effluent TSS (g·L⁻¹)</td>
<td>1.58</td>
<td>0.78</td>
<td>0.86</td>
</tr>
<tr>
<td>Final in-situ TSS (g·L⁻¹)</td>
<td>--</td>
<td>56.8 (8.34)</td>
<td>--</td>
</tr>
<tr>
<td>SRT (d)</td>
<td>--</td>
<td>251</td>
<td>--</td>
</tr>
<tr>
<td>Biogas Yield (L·L⁻¹·d⁻¹)</td>
<td>1.4</td>
<td>2.05</td>
<td>3.44</td>
</tr>
</tbody>
</table>

Sludge was washed and evaluated as outlined previously to observe any visual differences in granular sludge characteristics between the temperature levels (Figure 4-4). All three samples exhibited some degree of granulation with similar maximum granule sizes. The 35°C sludge appears to have a broader distribution of grain sizes. A review of theories of sludge granule development (Hulshoff Pol et al., 2004) indicates development of a nucleus which then grows into a mature granule by adding biomass. It is therefore hypothesized that good distribution of granule sizes may be an indicator of a healthy digester as this could indicate that new granules are being formed in and retained by the system. The 45°C and 55°C sludge samples show progressively less variety in size distribution, indicating that at the time the samples were taken, the digesters may have been washing out material. 55°C sludge is predominated by what appears to be residual material from the dairy waste seed sludge.
Figure 4.4: Washed granulated sludge recovered from IBR anaerobic digesters at conclusion of experiments. Photographs show sludge overlaid on a 2 mm grid for (L-R) 35° C, 45° C, and 55° C reactors.

The solids mass balance outlined in Equation 1 was used to calculate Solids Retention Time at peak OLR as reported in Table 4.3. Calculated SRTs of 254, 194, and 102 days were characteristic of retained biomass reactors and promising indicators of the suitability of the reactor design for treating high strength wastes. It must be noted, however, that the purpose of the design of the Induced Blanket Reactor is to treat high solids waste streams that will compete hydraulically for space in the reactor with the sludge granules critical to treatment.

Compounding the problem of retention, high solids substrates may also behave as relatively high viscosity fluids, impacting the settleability of solids. Care must therefore be taken in design of reactors to ensure that hydraulic considerations including fluid viscosity, upflow velocity, and particle settling characteristics are appropriately accounted for. These factors should be included as formal design criteria if the immobilized biomass/high SRT assumption is going to be applied, or a conservative assumption of HRT = SRT must be made to ensure that reactor performance is not impacted. It seems reasonable to assume that the 254 day SRT observed in the 35° C reactor may be indicative of a practical upper bound for SRT given the optimized feed, the absence of competing solids, and the minimal viscosity of the reactor fluid, and HRT = SRT may be assumed to be the lower bound for the system.

**SYSTEM PERFORMANCE**

It should be noted that differences in reactor performance as a function of temperature may not be ascribed to reactor geometry exclusively. The seed culture for the experiments was developed at
approximately 32°C. Adapting the same culture to the three temperature levels examined in these experiments requires raising the temperature of the culture 10-20°C. As temperature increases, cultures will adapt and evolve to the new environment so long as they remain viable. The process of selection must necessarily result in a reduction in culture diversity as temperature increases given the common gene pool that the system started with. Differences in system performance at different temperatures confound the biological effects of culture adaptation with the differences in heating rates and reaction rates. Similarities in reactor behavior across temperature ranges are therefore probably more significant than differences that cannot be ascribed to specific effects.

Figure 4-5 shows reactor performance at the three temperature levels and the three organic loading rates applied to the system. Steady state was reached for all nine treatments. Instances of operational inconsistency that impacted system performance included two system wide feed shutdowns, the first on days 76-78 to address problems with substrate mixing and the second from days 112 and 117 to fix a crack in the 55°C reactor wall. The 45°C reactor feed was also shut down from days 165-74 in response to a drop in system pH and decrease in biogas production and effluent quality.

Rapid system response to the absence of new substrate on shutdown and to the re-application of the feed on startup is readily observed as a function of biogas output which can be seen to drop when feed was shut off, and then recover almost immediately when feed was restarted. The reactor appears to rely on gas induced mixing to maintain contact between the sludge and the substrate in the liquid volume. Previous studies have shown that there is sufficient mixing due to heat flux through the reactor to maintain the liquid volume in a nearly completely mixed state, but the steep decline in biogas production on feed shutdown is indicative of the effective cessation of substrate reduction and importance of contact between substrate and biomass to treatment. Given that the liquid volume still contains approximately 3-15 g·L⁻¹COD (Table 4-4) at feed shutdown, more than sufficient to support continued activity, a lack of contact (mixing) between substrate and the sludge bed is indicated.
Figure 4-5: Loading rates, biogas yields, and effluent quality for reactors. SBY=Specific Biogas Yield. OLR eff = effluent organic loading rate and is a representation of effluent quality presented in terms of loading rate so as to be consistent with OLR inf, the influent organic loading rate. Gaps in OLR inf represent feed shutdowns as discussed in the text. Cyclic response in gas output is considered to be an function of feed strategy and acceptable within the bounds of steady state behavior.
Table 4-4: pH, solids, and COD mass balance parameters for IBRs operated at temperature levels indicated at steady state. CH₄ (%) reported is as calculated from mass balance at steady state. Standard deviations for sample sets reported in parentheses represent consistency of reactor performance as an indicator of spread of the measured distribution.

<table>
<thead>
<tr>
<th>Influent OLR (g COD·L⁻¹·d⁻¹)</th>
<th>35°C</th>
<th>45°C</th>
<th>45°C</th>
<th>55°C</th>
<th>55°C</th>
</tr>
</thead>
<tbody>
<tr>
<td>pH</td>
<td>7.6 (0.2)</td>
<td>7.7 (0.3)</td>
<td>7.8 (0.2)</td>
<td>7.8 (0.3)</td>
<td>7.8 (0.2)</td>
</tr>
<tr>
<td>Effluent COD (g·L⁻¹)</td>
<td>6.66 (3.84)</td>
<td>3.32 (1.14)</td>
<td>8.37 (2.25)</td>
<td>6.88 (3.54)</td>
<td>4.86 (1.33)</td>
</tr>
<tr>
<td>COD Removal (%)</td>
<td>78%</td>
<td>89%</td>
<td>73%</td>
<td>88%</td>
<td>84%</td>
</tr>
<tr>
<td>Residual OLR (Effluent) (g COD·L⁻¹·d⁻¹)</td>
<td>0.58 (0.33)</td>
<td>0.44 (0.15)</td>
<td>2.20 (0.59)</td>
<td>0.60 (0.31)</td>
<td>0.64 (0.17)</td>
</tr>
<tr>
<td>CH₄ (%)</td>
<td>57</td>
<td>66</td>
<td>64</td>
<td>61</td>
<td>58</td>
</tr>
<tr>
<td>Specific CH₄ Yield (g·L⁻¹·d⁻¹)</td>
<td>0.79 (0.15)</td>
<td>1.36 (0.07)</td>
<td>2.21 (0.43)</td>
<td>0.79 (0.22)</td>
<td>1.28 (0.14)</td>
</tr>
<tr>
<td>Specific Biogas COD (g COD·L⁻¹·d⁻¹)</td>
<td>2.09 (0.39)</td>
<td>3.58 (0.20)</td>
<td>5.84 (1.14)</td>
<td>2.09 (0.57)</td>
<td>3.38 (0.36)</td>
</tr>
<tr>
<td>VS (g·L⁻¹)</td>
<td>2.34 (1.87)</td>
<td>1.89 (2.83)</td>
<td>2.28 (0.67)</td>
<td>2.27 (1.54)</td>
<td>2.98 (4.22)</td>
</tr>
<tr>
<td>TS (g·L⁻¹)</td>
<td>10.6 (2.67)</td>
<td>9.8 (2.92)</td>
<td>10.1 (0.76)</td>
<td>10.1 (2.21)</td>
<td>10.7 (4.39)</td>
</tr>
<tr>
<td>FS (g·L⁻¹)</td>
<td>8.23 (1.08)</td>
<td>7.99 (0.48)</td>
<td>7.87 (0.14)</td>
<td>7.77 (0.80)</td>
<td>7.78 (0.43)</td>
</tr>
<tr>
<td>VS Removal (%)</td>
<td>82%</td>
<td>85%</td>
<td>82%</td>
<td>82%</td>
<td>77%</td>
</tr>
</tbody>
</table>

Effluent quality as concentration of COD improved for the 35°C and 45°C reactors between the low- and the mid-range loading rates, then decreased at the highest loading rate. The 55°C reactor effluent quality improved consistently as loading rate increased. Effluent COD concentrations show a significant difference in consistency as measured by the standard deviation of the distribution with the lower temperature reactors being superior in terms of producing a consistent effluent. The lowest loading rate exhibited the worst effluent consistency as measured by the standard deviation of the effluent COD loading distribution for all reactors. This may be a function of the increased mixing in the reactor induced by increased biogas production as discussed in Chapter 3. The increase of biogas production results in an increase in mixing energy input in the reactors, which then improves effluent quality by improving contacting and thus mass
transfer between the active solids and substrate rich liquid, and effluent consistency by homogenization of reactor contents.

In terms of COD removal, the 45° C reactor was the most effective at the highest loading rate, but the reactor was unstable, demonstrating a gradual reduction in pH from days 154-164 with a corresponding decrease in biogas production until reactor feed was cut off to attempt to stabilize the system. A residence time distribution study using Rhodamine WT dye (RWT) at 10 mg·L was being conducted on the reactor at this time. Since RWT was not inhibitory to the other reactors, it is not considered to be a likely cause of the upset, but it cannot be eliminated as a possible contributing factor. The reactor recovered to an extent in that the effluent and biogas stabilized when the reactor feed was restarted after 10 days at the previous OLR. The system was operated for another 20 days, but it did not approach previous treatment levels.

Methane production for the 35° and 45° C reactors, as calculated from the mass balance, was within the range of values expected for anaerobic digestion (57-69%). The 55° C reactor methane production was significantly lower, indicating inhibition of methanogenesis that appears to be similar to observations of ammonia inhibition in an un-acclimated culture at thermophilic temperature (Angelidaki et al., 1993; Hansen et al., 1998; Hashimoto, 1986). Ammonia data was not taken on the system, however, so this hypothesis remains untested and the precise cause of the significant difference in methane production between the 55° C system and the others is unknown.

Methane production was also measured directly using gas chromatography as described. Average values of 75%, 74%, and 72% were recorded for the 35°, 45°, and 55° C reactors at ultimate OLR and steady state. These values are regarded as unreliable. Tedlar® gas sampling bags were stored up to 4 weeks prior to analysis under the faulty assumption that Tedlar® gas sampling bags were appropriate for storage. Upon further review, Tedlar® was found to have significant coefficients of permeability for CO₂ (11.1), N₂ (0.25), and O₂ (3.2) (all cc·(100 in²·24 hr·atm·mily¹)) (DuPont, 1995). Biogas is inherently anoxic; this, coupled with the low levels of N₂ and high concentrations of CO₂ also characteristic of biogas creates a differential in partial pressures between the biogas and normal atmosphere. The partial pressure differential and the permeability of the Tedlar® membrane encourage diffusion across the bag where CO₂ diffuses out of the bag and O₂ and N₂ diffuse in as the systems equilibrate. Diffusion is also impacted by
the surface area to volume ratio of the bags; larger sample containers will have proportionally less surface area, reducing the potential magnitude of sample concentration. The GC results were therefore disregarded as unreliable. Lack of a standardized protocol for biogas sample storage in Tedlar bags is potentially a significant issue in investigations of anaerobic systems.

pH was consistently observed at the high end of the 6.5-8.2 range presented by Speece (1996) as optimal for anaerobic digestion. Volatile, fixed, and total solids are also reported for the effluent. Volatile solids removal correlates well with COD removal as expected.

COD removal capacity, expressed as % of influent COD removed is comparable to other high rate reactors at steady state (Speece, 1996; Stronach et al., 1986) in the 35°C and 45°C reactors with efficiencies ranging from 73-90%. Given the optimal feed conditions for the study (i.e. completely dissolved substrate, no refractory solids), and that the reactor is intended for use with high solids waste streams, these values might be upper limit efficiencies for the system, although further work should be undertaken to determine whether this is the case. COD removal in the 55°C reactor was much lower than anticipated. This may be due to the possible ammonia inhibition discussed previously.

Figure 4-6 provides a graphical representation of the COD mass balance on the systems at steady state (Table 4-2). The data were plotted as moving averages with the period equal to the reactor HRT. All the reactors showed reasonable stability at low loading rates (2.7 g COD•L⁻¹•d⁻¹), although the calculated methane content of the biogas was significantly lower at 55°C.

The plots emphasize the significant difference in effluent quality between the intermediate and ultimate loading rates as well as the cyclical nature of the methane generation rate at the high OLRs. It is also evident that performance in terms of effluent quality decreases with increased loading rates, but in terms of mass removal per unit volume, the highest loading rate is the most effective (Table 4-4). This may have implications for decisions regarding implementation of the technology, depending whether effluent quality or energy production is the primary driver in design criteria development.
Figure 4-6: Steady state mass balances for IBRs at different HRTs and temperature levels. Data is plotted as moving averages with period = 1 HRT. Contribution to specific system COD at steady state is plotted on the Y axis as a function of time. COD content of biogas was calculated in accordance with Equation 2.

Figure 4-7 provides a graphical comparison of effluent quality and biogas quality for each temperature level and loading rate. Biogas quality and effluent quality only appear to be correlated in the 35° reactor. The 35° and 45° C reactors both show relatively tight distributions in the effluent quality sample sets. This is believed to be significant in that the IBR is designed to pass solids; this design feature implies less
control of the effluent condition and the associated potential for large variations in effluent quality.

Variations are not evident in these data, and while these results do not preclude such a possibility, they do show that under favorable conditions, the reactor is capable of providing a consistent effluent stream.

![Graph showing OLR vs pH for different temperatures]

**Figure 4-7:** Average biogas production and effluent quality for each system at steady state. X-axis represents organic loading rates for treatments. Y-axis represents both Effluent Organic Loading Rate and Specific Methane Production in units as indicated in the Legend. Error bars represent one standard deviation.

Figure 4-8 shows the relative proportions of volatile fatty acids detected in the effluent and the system pH over the duration of the experiment. Effluent was monitored for acetic, propionic, butyric and valeric acids. Valeric acid was not detected in significant concentrations in any of the samples. Specce (2006) reports that propionate inhibition is not likely at concentrations below 3 g·L⁻¹, well above the levels detected at any point in this study.

pH was generally stable, indicating sufficient buffering capacity in the media to address any acidification with the exception of the previously mentioned drop in pH in the 45°C reactor, which unfortunately correlates with a gap in the VFA data for that system. It is also interesting to note that none of these treatments resulted in significant increases in VFA concentrations when OLR was doubled from 4.2 to 8.4 g COD·L⁻¹·d⁻¹ at Day 126 although a significant decline in VFA concentration in the 55°C system, and a less dramatic though still clear decline in VFA concentration in the 35°C system were observed. These trends could be interpreted to indicate that utilization of VFAs by methanogens was outstripping organic acid production. Given that acidogens are generally assumed to reproduce faster than methanogens when substrate is in excess this is a curious result. It could also be noted, however, that this
sudden decrease in VFAs is largely confined to the 55°C reactor, and that reactor was subject to a
significant process upset immediately prior to this shift in OLR as previously described; the results may not
be representative for a system that does not have similar issues of cooling, storage at a depressed
temperature, and reheating.

Figure 4-8: VFA concentration in effluent plotted with pH

For the 35°C reactor these data show very consistent and apparently serially correlated trends. There was
a significant trend indicating a decrease in total VFAs in the effluent between days 126 and 150,
corresponding to the time when the reactor was adapting to the final OLR. At approximately day 159, VFA concentrations bounced back. Propionate was the predominant acid species in the system. The 45°C and 55°C reactors also showed serial correlation in VFA concentrations. For the 45°C system, acetate was the dominant species, while for the 55°C reactor, acetate was higher until the decline in total acid concentration when propionate overtook acetate, retaining that spot when the VFA concentrations stabilized towards the end of the study.

CONCLUSIONS

Three Induced Bed Reactor anaerobic digesters were operated successfully for 194 days at three temperature levels (35°C, 45°C, and 55°C) on a completely dissolved substrate feed at maximum OLR of 8.4 g COD·L⁻¹·d⁻¹ with a minimum HRT of 3.8 days. The most robust system, as indicated by its ability to adapt to new loading rates with a minimum of disruption to effluent and biogas quality was the 35°C reactor.

All three reactors developed beds of granulated, stratified on the bottom of the reactors. The granules were interspersed with thin rod-shaped structures that had the appearance of clipped hairs 1-3 mm in length. Bed formation demonstrated that the reactor feed systems, substrate, loading rates, and flow rates utilized for the experiments were appropriate for developing a dense granular sludge.

Settled sludge volumes (SSV) as determined using Imhoff cones were consistent with observations made through reactor walls; SSV for the 35°C reactor corresponded to a sludge bed occupying approximately half of the reactor volume; for the 45°C reactor, the sludge bed occupied approximately 25% of the reactor volume. The 55°C sludge bed SSV corresponded to approximately 12% of the reactor volume.

Flocculent sludge was also observed to accumulate in the upper chamber of the reactors above the septum. Spikes in effluent suspended solids and COD would occur periodically when these flocculated sludge beds would apparently reach a critical volume and mix with the effluent impacting both effluent quality and consistency.

Calculated SRTs of 254, 194, and 102 days were characteristic of retained biomass reactors and promising indicators of the suitability of the reactor design for treating high strength wastes. For purposes
of design, in the absence of more permissive data the 254 day SRT observed in the 35° C reactor may be conservatively assumed to be indicative of a practical upper bound for SRT and HRT = SRT may be assumed to be the lower bound for the system SRT.

The IBR appears to rely on gas induced mixing to maintain contact between the sludge and the substrate in the liquid volume. Previous studies have shown that there is sufficient mixing due to heat flux through the reactor to maintain the liquid volume in a nearly completely mixed state, but the steep decline in biogas production on feed shutdown is indicative of the effective cessation of substrate reduction and the importance of contact between substrate and biomass to substrate reduction. An effective means of enhancing reactor performance might be to recirculate digester gas through the sludge bed to enhance mixing of settled sludge and the liquid fraction of the reactor contents.

COD removal capacity is comparable to other high rate reactors at steady in the 35° C and 45° C reactors with observed efficiencies ranging from 73-90%. COD removal in the 55° C reactor was much lower than anticipated. The 45° C reactor was the most effective in terms of substrate reduction at the highest loading rate, but the reactor was unstable demonstrating a gradual decline in system pH necessitating a feed shutdown to permit the system to recover. Effluent COD concentrations show a significant difference in consistency as measured by the standard deviation of the distribution with the lower temperature reactors being superior in terms of producing a consistent effluent. The lowest loading rate exhibited the worst effluent for all reactors.

Methane production for the 35° and 45° C reactors, as calculated from the mass balance, was within the range of values expected for anaerobic digestion (57-69% of biogas by volume). The 55° C reactor methane production was significantly lower, possibly due to the effects of toxicity inhibition.

pH was generally stable, indicating sufficient alkalinity in the substrate to buffer any organic acid loading encountered. It was consistently observed at the high end of the 6.5-8.2 range presented by Specce (1996) as optimal for anaerobic digestion.

Reactor performance in terms of minimizing effluent COD decreases as loading rates are increased, but in terms of mass removal per unit volume, the highest loading rate is the most effective; the faster the substrate is run through the system, the more COD is converted even though effluent quality is worse.
Results of this study should be interpreted with care; performance data is a function of both reactor design and biological characteristics of the cultures used to operate the systems. No attempt was made to optimize the biological characteristics of the cultures for the experiments, and results should not therefore be considered as indicators of optimal performance of the IBR at a given temperature. The study demonstrates that the IBR can be operated successfully at a range of temperatures. Issues with system stability in the 45° and 55° C reactors may be attributable to the reactors themselves, the starter culture used to develop the operating cultures, the evolved operating cultures, or a combination of factors.

**Implications for Design**

Care must be taken in design of reactors to ensure that hydraulic considerations including fluid viscosity, upflow velocity, and particle settling characteristics are appropriately accounted for. These factors should be included as formal design criteria if the immobilized biomass/high SRT assumption is going to be applied, or a conservative assumption of HRT = SRT must be made to ensure that reactor performance is not impacted.

Mixing of the upper chamber of the IBR could be considered to enhance effluent quality and consistency given the tendency of this upper volume to accumulate and release flocculent sludge. Mixing of the lower reactor volume by recirculating and sparging biogas through the sludge bed to enhance contacting should also be considered.

**REFERENCES**


CHAPTER 5

THERMOPHILIC ANAEROBIC DIGESTION USING AN INDUCED BED REACTOR

The Induced Bed Reactor (IBR) was developed at Utah State University (USU) to apply high-rate anaerobic digestion techniques to highly-suspended solids content substrates (6-12% total solids) such as food waste and dairy manures. This technology has been successfully implemented at full-scale multiple installations in the United States and Canada as a waste treatment and energy production technology.

A 58L bench scale reactor was operated at 55°C with three organic loading rates and three corresponding hydraulic retention times using a dextrose/yeast extract substrate at 30.6 g/L COD. Influent and effluent streams were monitored for solids composition, gas quality and quantity, and chemical oxygen demand (COD). Residence time distribution (RTD) was determined using step input tracer studies with Rhodamine WT and Lithium tracers.

The IBR was successfully operated at thermophilic temperatures for 180 days, demonstrating a COD removal rate of 64% and a robust thermophilic culture. Under the conditions studied, the IBR exhibits complete mix behavior with no mechanical energy input. Development of a granulated sludge bed comprising a settled sludge volume of 120mL/L was evident. Further work should be done to determine conditions required for improved solids retention and thus improved performance.

INTRODUCTION

The purpose of anaerobic digestion in wastewater treatment is to reduce waste sludge volume and activity by fermenting the waste in the absence of oxygen. Anaerobic digestion relies on the symbiotic relationship between two general classes of anaerobic microorganisms to catabolize carbonaceous substrates to the relatively stable and innocuous end products of biosolids and biogas (Bryant et al., 1967).

The relatively long doubling times of anaerobes require either long residence times (20-40 d) to allow populations to accumulate to levels permitting significant treatment or separation of hydraulic retention time (HRT) and solids residence time (SRT) to permit high-rate treatment. Methods for accomplishing this include separating the active solids from the effluent and recycling them to the digester, by directly

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1 Coauthored by J. S. Dustin and C. L. Hansen
utilizing an attached growth process, or through the use of an induced bed or blanket of active biosolids that self-select to remain in the reactor as the substrate passes through (Gerardi, 2003; Lettinga et al., 1980; McCarty, 1981; Tchobanoglous et al., 2003). High rate processes like these generally require a relatively low suspended solids concentration in the influent to avoid dilution of the active biomass and to avoid operational problems in the digester (plugging, sedimentation, short circuiting, etc).

The inability to treat wastes with high influent suspended solids content poses a problem application of high rate processes to substrates like those associated with food processing wastes and livestock manures. In the early 1990s, Dr. Conly Hansen at Utah State University proposed a digester configuration, the Induced Bed Reactor (IBR), that was designed to address the operational limitations of applying high rate treatment techniques to high strength, high solids substrates.

Anaerobic digestion has been extensively studied at temperatures ranging from psychrophilic to thermophilic, and thermophilic digestion has demonstrated some advantageous properties. The van’t Hoff-Arrhenius equation predicts that the product formation rate will approximately double for every 10°C increase in reaction temperature. The implication of this relationship is that required reactor volume is inversely proportional to temperature (Lettinga et al., 1980; Tchobanoglous et al., 2003). Additionally, human pathogens are mesophiles. Exposure to thermophilic temperatures disrupts microbe cell walls and metabolic function and provides effective pathogen reduction.

Varel et al. (1977) found that a functional thermophilic (55°C) seed culture could be developed from raw cattle waste over an eight day incubation period. The same culture was later adapted to 60°C. Utilizing this culture in a semi batch process, he was able to generate and maintain a thermophilic culture capable of degrading a raw manure feed. Subsequent work (Varel et al., 1980) considered reactors maintained at different temperatures from 30 to 60°C in 5-degree increments. They observed an approximately fourfold increase in biogas production for a reactor at 60°C over one at 40°C. It was also shown that higher temperature reactors could maintain proportionally higher substrate loading rates.

Mackie and Bryant (1995) also looked at fed batch thermophilic (60°C) and mesophilic (40°C) digestion of cattle waste. They demonstrated higher energy content in the thermophilic biogas and higher reaction
rates in the thermophilic systems. They also observed that at high loading rates, thermophilic digesters converted substrate more efficiently than mesophilic systems.

The IBR relies on biomass retention in the digester to provide an accumulation of active solids, similar to the operational principles first proposed by Lettinga et al. (1980) for UASB reactor design (Figure 5-1). UASBs operate on the principle of separation of SRT and HRT via granulated biomass retention. The active biomass acts as an effective filter, metabolizing available carbon to CO$_2$ and CH$_4$. In the UASB, influent enters the reactor from the bottom via a diffuser to provide equal distribution through the sludge bed and minimize short-circuiting. Substrate transits the sludge bed vertically, converting chemical oxygen demand (COD) to biogas. Biogas is directed via the deflector baffles to the 3-phase separator, while treated liquid effluent flows around the baffles and the separator and over a weir to the reactor outlet.

Figure 5-1: Cutaway view of an Upflow Anaerobic Sludge Blanket (UASB) reactor (L) and an Induced Bed Reactor, IBR (R).

In 1993, Dr. Hansen began working with digestion of high-suspended solids, and high strength wastewaters in reactors configured as UASBs in an attempt to translate some of the advantages of the high-rate small footprint systems to agricultural and food wastes. This resulted in the development of the
Induced Bed Reactor (IBR) (Figure 5-1), an upflow design that preserves some of the biomass concentrating characteristics of the UASB while allowing for the processing of complex high solids and wastewaters such as dairy manures. Note the septum baffle which separates the mixed liquid volume from the headspace, the diffuser plate which provides substrate diffusion into the sludge bed, and the auger which keeps the gas outlet in the septum baffle clear of solids. Liquid effluent exits the reactor via an inverted trap above the septum.

The differences between the two reactor types are derived from the need to address the high solids content of the influent. Inlet and outlet conditions and the gas separation mechanism were modified to prevent plugging in the reactor. The HRT for the IBR is 4-8 times that required for the UASB. The principal advantages of the IBR system are that the system a) reduces the need for solids separation over what would be required for other high-rate systems, b) with a 4-day HRT requires 5x less volume than a similarly performing plug or complete mix digester with a 20-d HRT, and c) permits the thermal processing of suspended solids in the influent at thermophilic temperatures. Although recalcitrant suspended solids in the influent may not be subject to digestion, they can pass through the reactor where they are raised to thermophilic temperatures and gain the benefit of pathogen destruction before being carried out with the effluent stream.

While the IBR has been successfully implemented with over 4,000 m$^3$ of installed capacity in the US and Canada, there is no published data on operation of the system at temperatures above 35°C. The purposes of this study were to observe the IBR at thermophilic temperatures and gather data on digestion (operational behaviors, conversion performance, and biomass characteristics), and mixing characteristics (residence time distribution study).

**MATERIALS AND METHODS**

**DIGESTION**

*Experimental Apparatus*

The experimental apparatus for the research is shown in Figure 5-2. A 58 L working volume bioreactor was constructed from extruded acrylic tubing 300 mm ID and 914 mm high. Influent feed was controlled by peristaltic pumps (Masterflex 7523, Cole-Parmer, Inc., Vernon Hills, IL). During the experiments, the
system was maintained at 55°C by heat tape applied to the bottom 300 mm of the reactor and controlled by a digital temperature switch (Love TS-13011, Dwyer Instruments, Michigan City, IN) utilizing a thermistor placed in a thermowell at the mid-height of the reactor for feedback. The reactors were insulated with 90 mm of fiberglass insulation. Gas production was measured via a thermal mass flow meter (822S-L2-ON1-PV1-V4, Sierra Instruments, Monterey, CA).

![Diagram](image)

**Figure 5-2: Schematic design of the research apparatus**

Substrate was refreshed in the feed tank at 7-10 day intervals, and maintained in a mixed state using a variable speed stand mixer operated at 1800 RPM (EVP1P25MIC48, Lightnin, Rochester, NY). The feed tank was maintained in a cold room at 5°C to preserve substrate quality. The feed pump was set to provide a constant rate feed to the digester at the desired HRT. Foam control was provided by a semi-continuous injection of Sigma Antifoam B (Sigma Aldrich, Inc. St Louis, MO) via a diaphragm metering pump (LMI AA97, Milton Roy Americas, Ivyleand, PA) to maintain an antifoam concentration in the headspace of 30
mg/L. Effluent gas was routed through a liquid capture foam trap, thence to a refrigerated water column where excess water vapor was removed by bubbling the gas through the column at 5°C before warming the gas to 22°C for flow measurement.

Influent samples were collected in the substrate feed tank. Gas samples were collected in Tedlar bags (Fisher Scientific) for composition analysis, with excess gas vented via a fume hood. Effluent samples were collected at the liquid effluent port. Composite samples for analysis of reactor contents at shutdown were collected as grab samples by first mixing the reactor volume then taking a representative sample from the water column with a sludge sampler (Sludge Judge, Nasco, Fort Atkinson, WI).

**Substrate Composition**

The substrate solution had a COD of 30,600 mg/L and was composed of 23,438 mg/L of dextrose, plus the following nutrients (mg/L): 2,500 yeast extract, 2,656 NH₄Cl, 525 K₂HPO₄, 225 FeCl₂·4H₂O, 469 CaCl₂·2H₂O, 391 MgSO₄·7H₂O, and 313 KCl. To prevent required microbial trace element deficiency, a trace nutrient solution (500mg/L H₂BO₃, 500 mg/L ZnCl₂, 300 mg/L CuCl₂, 5000 mg/L MnSO₄·H₂O, 500 mg/L (NH₄)₆Mo₇O₂₄·4H₂O, 500 mg/L AlCl₃, 500 mg/L CoCl₂·6H₂O, and 500 mg/L NiCl₂) was added by 0.01% (v/v) to each batch of substrate. NaHCO₃, 10,000 mg/L, was added to maintain initial buffering capacity, and tap water (City of Logan, UT) was used as dilution water. The components were similar to those used for cultivating anaerobic bacteria in an ASBR by Cheong et al. (2007).

**Culture Development**

The thermophilic culture was developed from seed sludge taken from an operating IBR (Wade Dairy, Ogden, UT). The seed sludge was passed through a ¼” mesh screen to remove larger solids. The reactor was filled with a mixture of 1/3 sludge, 1/3 water, and 1/3 substrate, brought to temperature (55°C) over 5 days, and allowed to operate in batch until the bioavailable energy value of the substrate was effectively exhausted as measured by the absence of biogas production.

Mesophilic IBRs are typically operated at HRTs of 3.5-4 days. For purposes of comparison with other ongoing laboratory studies, the target operational point for this reactor was an HRT of 3.8 days with a
corresponding organic loading rate (OLR) of 8.0 g COD/(L·d). This was reached by increasing the OLR in steps as outlined in Table 5-1.

<table>
<thead>
<tr>
<th>Time (days)</th>
<th>OLR (g COD/(L·d))</th>
<th>HRT (days)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-50</td>
<td>0</td>
<td>--</td>
</tr>
<tr>
<td>51-95</td>
<td>2.7</td>
<td>11.5</td>
</tr>
<tr>
<td>95-126</td>
<td>4.0</td>
<td>7.6</td>
</tr>
<tr>
<td>126-194</td>
<td>8.0</td>
<td>3.8</td>
</tr>
</tbody>
</table>

Monitoring and Analysis

During the experiments, the reactor was continuously monitored for gas production. Effluent grab samples were taken daily and analyzed for total solids (TS), volatile solids (VS), fixed solids (FS), and total COD. Solids concentrations were measured in accordance with Standard Methods (APHA-AWWA-WEF, 2005). COD was measured by the closed reflux colorimetric method and settled sludge volume was determined using an Imhoff Cone (APHA-AWWA-WEF, 2005).

Biogas flow rates were recorded at 6 Hz using LabView 8.2 software (National Instruments, Austin, TX). The methane content in the biogas was analyzed by gas chromatography (GC, HP 6890 series, Hewlett-Packard, Wilmington, DE) using a capillary column (RT-Msieve 5A PLOT, Restek, Bellefonte, PA) with dimensions of 30.0 m x 320 μm x 30.0 μm. The column temperature was 35°C, while the inlet port and thermal conductivity detector temperatures were 50 and 200°C, respectively. Argon was used as the carrier gas at a flow rate of 1.5 mL/min. Gas standards were obtained from Scott Specialty Gases (Plumsteadville, PA) for calibration.

Mixing

Experimental Design

Mixing behavior in reactors is investigated through the use of Residence Time Distribution (RTD) studies wherein a conservative tracer is injected into a reaction vessel and observed as it exits. The resulting observations can be analyzed and compared to standard curves to determine degree of mixing in the system. For this work, two sets of experiments were conducted (Table 5-2). The first was a control
study in clean water. The second set incorporated the effects of active biomass and digestion on mixing in the system at steady state. Tracers were mixed into the digester feed tank at known concentrations, and fed via the influent feed pump as a step function (viz. concentration in feed was switched from 0% to 100% instantaneously and maintained at 100% until the effluent dye concentration equaled the influent dye concentration).

Table 5-2: RTD study conditions

<table>
<thead>
<tr>
<th>Description</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>Clean Water</td>
<td>RWT\textsuperscript{a}</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Active Digestion</td>
<td>RWT\textsuperscript{a}</td>
<td>Li\textsuperscript{+}\textsuperscript{b}</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>2</td>
<td></td>
<td>Washout</td>
<td></td>
</tr>
<tr>
<td>[Tracer], mg L\textsuperscript{-1}</td>
<td>1.5</td>
<td>1.5</td>
<td>1.0</td>
<td>10</td>
</tr>
<tr>
<td>COD\textsubscript{in} g L\textsuperscript{-1}</td>
<td>--</td>
<td></td>
<td>30.6 (2.1)</td>
<td></td>
</tr>
<tr>
<td>OLR\textsubscript{in} g COD\textsuperscript{-1} L\textsuperscript{-1} d\textsuperscript{-1}</td>
<td>--</td>
<td></td>
<td>8.05 (0.9)</td>
<td></td>
</tr>
<tr>
<td>HRT (days)</td>
<td></td>
<td></td>
<td></td>
<td>3.8</td>
</tr>
</tbody>
</table>

\textsuperscript{a} RWT = Rhodamine WT (Kingscote Chemicals, Miamisburg, OH)
\textsuperscript{b} Li\textsuperscript{+} = Lithium ion from Lithium Chloride (Sigma Aldrich, St Louis, MO)

Experimental Apparatus

Experimental apparatus was the same as that used for the digestion experiments. Tracer samples were collected at the digester effluent port at recorded time intervals and comprised both mixing cup and grab samples. Grab samples represent a discrete interval in the tracer history and are taken by simply capturing the effluent at a specific time period. Mixing cup samples were obtained by collecting the full volume of the reactor discharge over a known period of time, then mixing the aggregated sample and taking an aliquot for analysis. Results of mixing cup samples yield an average effluent tracer concentration over the sample aggregation period. They can be incorporated into the grab sample dataset by recording the average time corresponding to the sample interval.

Tracer

Tracers selected were rhodamine WT (RWT) (Kingscote Chemicals, Miamisburg, OH) and lithium ion from lithium chloride (Li\textsuperscript{+}) (Sigma Aldrich, St Louis, MO). RWT was selected as the primary tracer due to cost of material and cost of analysis, with a confirmatory study was performed using Li\textsuperscript{+} to provide a mass
balance and to confirm the qualitative results obtained from the RWT study given that rhodamine WT is known to have two isomers, one of which exhibits sorptive behaviors which have the potential to impact study results (Vasudevan et al., 2001). Lithium also has disadvantages including the potential for toxicity in anaerobic systems (Anderson et al., 1991). These effects are confounded with other variables in the study and cannot be separated, but both chemicals are widely regarded as appropriate for digester RTD modeling in spite of their limitations (Denbigh and Turner, 1965; Leighton and Forster, 1996; Lou et al., 2006; Tchobanoglous et al., 2003).

**Monitoring and Sample Analysis**

RWT samples were analyzed by pipetting well mixed samples into 96 well black opaque plates (Corning Costar) and analyzing the samples in a plate reader fluorimeter (S4MLFPT, BioTek, Winooski, VT) with Gen 5 software (v1.04.5). The tungsten lamp was used with an excitation wavelength of 530 nm, and an emission wavelength of 585 nm. Sensitivity was set to 100, with a top probe vertical offset of 4.0 mm and a column offset of 0 mm. Lithium tracer samples were analyzed by the Utah Veterinary Diagnostic Laboratory (Utah State University, Logan, UT) using internal standard operating procedure 1245.0 for ICP/MS.

RWT can be difficult to analyze given the background fluorescence of the effluent matrix and the variability of effluent solids composition (and thus fluorescence). The selection of the step input function in the design of the study permits a self-referencing two point calibration for each treatment where the effluent fluorescence immediately prior to beginning tracer injection is assumed to correspond to a tracer concentration of zero, and the maximum effluent fluorescence observed at the end of the study is assumed to correspond to a tracer concentration of 100%. The result is that while the RWT studies are valid for determining qualitative mixing behavior, since they cannot be referenced to an external standard, they cannot be used to verify a mass balance on the reactor and therefore can’t be used to quantify such characteristics as dead zones and bypass/short circuiting flows (Fogler, 2006; Levenspiel, 1999). Given the absence of Li+ in the effluent matrix background, the Li+ results are independently calibrated to external standards and can be used to confirm the results of the RWT studies and to complete a mass balance on the mixed contents of the system.
RTD data was analyzed in accordance with the methods proposed by Danckwerts (1953). Data was resolved to normalized cumulative residence time distribution curves (F-curves) by plotting the observed tracer concentration normalized to maximum tracer concentration, \( F(t) \), vs. the observed time normalized to the hydraulic retention time, \( \theta \). The theoretical hydraulic retention time (HRT) for the system, \( \tau \), is calculated from the liquid volume of the reactor divided by the influent flow rate. \( \theta \) is the normalized time, \( t/\tau \) (unitless). \( F(\theta) \) is the normalized tracer concentration calculated from concentration of tracer detected in the effluent divided by maximum tracer concentration, \( C_0 \) (unitless). The normalized curves can be directly compared to idealized curves for interpretation of data (Danckwerts, 1953; Fogler, 2006; Levenspiel, 1993, 1999; Wen and Fan, 1975).

The F-curve can be analyzed as a cumulative statistical distribution where the variance of the observed values indicates the spread of the distribution; the larger the variance, and thus the spread of the RTD, the higher degree of mixing. If the variance is normalized with respect to time, it can also be related to the dispersion number. The dispersion number is the inverse of the Peclet number, and serves as an estimate of the axial dispersion in a reactor which can be correlated to degree of mixing (Tchobanogloss et al., 2003; USEPA, 1986). A value above 0.1 is regarded as high dispersion, and an indicator of a well mixed reactor. A value of 1 indicates a completely mixed system and conformance to an ideal CSTR.

RESULTS AND DISCUSSION

BIOLOGICAL PERFORMANCE

Figure 5-3 shows biogas generation as a function of OLR for the duration of the study. The period from 0-50 days indicates the initial loading and acclimatization of the active biomass to the reactor conditions. At day 50, steady state reactor feed commenced in accordance with the schedule outlined in Table 5-1. The process feed was shut down from day 112 to 116 when the reactor wall cracked creating a slow leak and necessitating a repair. The reactor contents were salvaged and stored at 4°C for two days until the reactor could be sealed again. The reactor contents were replaced and the reactor was heated back to operating temperature (55°C) over 48 hours. Although the gas production at an average 89 L/d (1.5 L·L−1·d−1, )
seemed to have recovered to the levels seen immediately prior to the shutdown, it was 26% lower than the 120 L·d⁻¹ (2.1 L·L⁻¹·d⁻¹) observed between days 92 and 101. It is possible that the crack in the reactor began to impact the performance of the system as early as day 101 as evidenced by the steep decline in gas production at that time. It is also possible that the shutdown had a negative impact on the reactor cultures, but the reactor appeared to recover as evidenced by the continued functioning of the digester at the levels observed immediately prior to the shutdown. Another feed shutdown occurred at day 131 when a blockage was cleared from the feed line.

![Figure 5-3: Biogas generation and organic loading rate](image)

The digester reached steady state for the target conditions (HRT = 3.8 days, OLR = 8.0 g COD·L⁻¹·d⁻¹) at 153 days of operation. Given an influent [COD] of 30.6 g·L⁻¹ and an effluent [COD] of 11.1 g·L⁻¹ (σ = 2.2 g·L⁻¹), the process demonstrated a consistent COD reduction of 19.5 mg/L or 64% (Table 5-3). Table 5-3 summarizes the average values for COD and VS destruction in the reactor at a 3.8 day HRT. Biogas production was measured at an average of 245 L/d, yielding a specific biogas production rate of 4.2 L biogas·L⁻¹·d⁻¹. Methane content of the biogas averaged 72% (σ =10%).

Table 5-3: Substrate depletion at steady state, OLR = 8.0 g COD/(L·d), HRT = 3.8 d

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Influent</th>
<th>Effluent</th>
<th>% Reduction</th>
</tr>
</thead>
<tbody>
<tr>
<td>OLR (g COD/(L·d))</td>
<td>8.0</td>
<td>2.7</td>
<td>66%</td>
</tr>
<tr>
<td>COD (g COD/L)</td>
<td>30.6</td>
<td>11.1</td>
<td>64%</td>
</tr>
<tr>
<td>VS (%)</td>
<td>1.28</td>
<td>0.36</td>
<td>72%</td>
</tr>
<tr>
<td>FS (%)</td>
<td>0.88</td>
<td>0.78</td>
<td>11%</td>
</tr>
</tbody>
</table>

When considered as a function of OLR, the reactor performance improves in terms of substrate conversion and effluent stability as the organic loading rate increases (Figure 5-4). At the two lower loading rates, there was very little difference in effluent quality as expressed as a function of OLR (Table 5-4). At the highest loading rate, the effluent quality decreased by a factor of two over that observed for the lower loading rates, but the substrate conversion increased by a factor of four. The distribution of effluent quality values stabilizes at the highest loading rate as evidenced by the significant reduction in the standard deviation of the measured effluent OLR.

Figure 5-4: IBR performance at steady state for various loading rates as a function of OLR.

At the conclusion of the study, the reactors were drained and the contents recovered. A well attached but very thin black biofilm was observed on the reactor walls. The sludge bed in the bottom of the reactor was an interlocked mat of fibrous material and granulated sludge (Figure 5-5). Settleable solids content of
the sludge was measured in an Imhoff Cone as 120 mL/L, corresponding to a sludge bed that would have occupied approximately 1/8 of the reactor volume.

Table 5-4: Effluent OLR comparison

<table>
<thead>
<tr>
<th></th>
<th>Average OLR (g sCOD·L⁻¹·d⁻¹)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Influent</td>
<td>2.7</td>
</tr>
<tr>
<td>Effluent</td>
<td>1.3</td>
</tr>
<tr>
<td>Substrate Conversion</td>
<td>1.4</td>
</tr>
<tr>
<td>Substrate Conversion (%)</td>
<td>51%</td>
</tr>
<tr>
<td>σ effluent</td>
<td>0.56</td>
</tr>
</tbody>
</table>

Figure 5-5: Photograph of washed sludge solids from IBR operated at 55°C. 2mm grid shown for reference.

RESIDENCE TIME DISTRIBUTION

Normalized cumulative residence time distribution (RTD) curves are shown for the reactor in Figure 5-6. An ideal completely stirred tank reactor (CSTR) curve is shown for comparison. From the graph it is evident that the fluid portion of the reactor behaves as a complete mix system with some arbitrary flow which causes the observed fluctuations about the ideal CSTR.

The RTD for the clean water (no solids) reactor shows complete mix behavior with a small lag which may indicate some plug flow behavior. This is believed to be an artifact of the heating mechanism as in observing the dye input to the reactor during the initial injection sequence, dye accumulated in the bottom 2% of the reactor, possibly indicating a thermocline in the region below the heat source. Once the injection
fluid, which entered the reactor at 4°C, warmed up sufficiently to interact with the circulating fluid above, it was incorporated into the reactor volume and mixed freely.

![Figure 5-6: Cumulative Residence Time Distribution (F-curves) for the IBR at 55°C. The five curves shown represent the four studies outlined in Table 2 and an idealized CSTR with perfect mixing. Clean Water (1) represents a control study using clean water and a Rhodamine WT (RWT) tracer with energy input limited to reactor heating. RWT(2), Li+(3), and RWT(4) represent results of active digestion studies wherein biogas evolution and the presence of solids in the reactor impacted mixing behavior.](image)

When the reactors are studied under active digestion conditions, the energy input to the system includes the baseline input from the thermal energy entering the system to maintain the reactor at operating temperature and the kinetic energy of the biogas bubbles formed in the sludge bed. The heat energy from temperature maintenance converts in part to kinetic energy as fluid rises and falls in response to density and viscosity gradients in the reactor. This is postulated to be the source of mixing forces in the clean water tracer studies. When the active digestion component is added, biogas is constantly being evolved in and released from the sludge bed. As these biogas bubbles transit the reactor wet volume, they induce shear forces in the digester contents, adding to the mixing of the thermally induced shear forces. The additional energy input from biogas generation appears to be sufficient to mix the liquid volume as demonstrated by the elimination of the plug flow tail evident in the clean water F-curve from the RWT and Li+ RTD F-curves.
With the incorporation of active solids and available substrate, the RTD curves for the lithium and rhodamine studies track the ideal CSTR well. The lithium data shows instances that appear to be consistent with some degree of short circuiting as exhibited by values for τ in excess of those predicted by the ideal CSTR model, but the oscillation around the ideal coupled with under prediction of τ for the rhodamine studies, indicates that this may be an artifact of the tracer. The geometry of the reactor, including the placement of the inlet and outlet, the sludge bed, and the location of two baffles between the inlet and outlet make the establishment of a consistent bypass stream unlikely under the study conditions.

For these data, the dispersion numbers all show complete mix behavior (Table 5-5). As would be anticipated, degree of mixing is clearly enhanced by the impact of the addition of active biomass and substrate with the dispersion approximately doubling. This is likely due to the mixing from the evolution of biogas in the sludge bed, and the shear induced as it transits the reactor.

Table 5-5: Calculated parameters derived from RTD studies for the IBR at 55°C, OLR = 8.0 g COD/(L-d)

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Clean Rhodamine</th>
<th>Active Rhodamine</th>
<th>Active Lithium</th>
</tr>
</thead>
<tbody>
<tr>
<td>τ, d (Vol/flow rate)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Variance, ( \sigma^2_{\text{Rh}} )</td>
<td>9.1</td>
<td>16.4</td>
<td>22.1</td>
</tr>
<tr>
<td>Normalized Variance, ( \sigma^2_{\text{NR}} = \sigma^2_{\text{Rh}} / \tau^2 )</td>
<td>0.63</td>
<td>1.14</td>
<td>1.54</td>
</tr>
<tr>
<td>Dispersion ( d = \sigma^2_{\text{NR}} )</td>
<td>0.32</td>
<td>0.57</td>
<td>0.77</td>
</tr>
<tr>
<td>Mixing</td>
<td>Complete mix</td>
<td>Complete mix</td>
<td>Complete mix</td>
</tr>
</tbody>
</table>

It should be noted that the mixing parameters derived by this study are necessarily limited to the behavior of the liquid fraction of the reactor contents. By definition, as a retained biomass reactor, the IBR maintains a separation between the solids retention time (SRT) and hydraulic (liquid) retention time (HRT) in the reactor. This requires that mixing energy be maintained below the threshold that would provide for uniform distribution of solids in the reactor and thus eliminate the difference between SRT and HRT.

CONCLUSIONS

The Induced Bed Reactor was operated successfully at thermophilic temperatures. Operational difficulties including equipment failure may have impacted development of an optimal microbial
community, but the reactor was operated successfully for over 180 days at 55°C. Equipment failure, subsequent sludge activity recovery, and restart of the reactor demonstrated the robustness of the culture developed in the experiment.

The IBR demonstrated a reactor efficiency of 64% in terms of waste strength reduction at an OLR of 8.0 g COD·L⁻¹·d⁻¹ and an HRT of 3.8 days. Effluent quality was highly variable at lower OLRS and HRTs, but stabilized when OLR was increased to the final steady state value.

The settled sludge volume (120 mL/L) was lower than that observed in IBRs operated at lower temperatures. The sludge appeared to be a combination of residual material from the seed sludge and some granulated sludge. A strongly granulated sludge was developed in reactors operated at 35 and 45°C under otherwise similar conditions. Further research is required to determine whether this is possible at higher temperatures with the IBR; the lower settled sludge volume may be an indication of an underloaded reactor.

The RTD analysis shows that the liquid fraction of the IBR operates as a complete mix reactor. The clean water tracer study represented the minimum energy input to the system and demonstrated that the heat flux creates sufficient fluid movement in the reactor to mix the contents over the HRT studied. Subsequent studies in active digesters added energy inputs due to gas evolution and evacuation as the biogas bubbles transited the column, and increased liquid mixing.

The IBR can be operated at thermophilic temperatures with the substrate used in these experiments, Further work should be done to determine conditions required for improved solids retention and thus improved performance.

References


CHAPTER 6
GENERAL SUMMARY

CONCLUSIONS

The following conclusions summarize the major findings of this research:

INTRODUCTION TO THE IBR

1) The IBR has been implemented in the United States and Canada with reasonable treatment success. The reactor appears to be capable of treating relatively high solids substrates such as the dairy manures considered herein with calculated influent TS concentrations ranging from 3.9-10.4%. It appears that relatively higher solids loading rates can be correlated to higher specific biogas production without reactor failure for the conditions observed, although it must be recognized that observations reported herein are derived from self-reported data, and should therefore be regarded with caution.

HYDRODYNAMIC MODELING

1) Lithium and Rhodamine WT were used to investigate the mixing behavior of the Induced Bed Reactor (IBR) anaerobic digester at laboratory (58 L) scale. Step input residence time distribution (RTD) studies in both clean water and active digesters showed that liquid fraction of the IBR operates as a complete mix reactor with potential for deviation from ideal CSTR behavior when analyzed using both the combined and CSTR in series models.

2) Energy input in the lab scale clean water tracer studies was limited to the heat required to maintain temperature in an insulated reactor. This represented the minimum energy input to the system and demonstrated that the heat flux creates sufficient fluid movement in the reactor to mix the contents over the HRT studied with a plug flow component equal to approximately 10% of the reactor volume. The clean water study shows no important difference in mixing as a function of reactor temperature.

3) Based on the results of the clean water RTD studies, a real CSTR compartment model with elements of bypass and dead space was selected for further investigation. This model was applied
to the tracer data along with the CSTR in series model, and nonlinear modeling demonstrated a
good fit with the data indicating CSTR behavior. The conclusion of the CSTRs in series model is
similar in that it considers the IBR to operate as 1.2 CSTRs in series.

4) Although rhodamine WT is widely used in digester modeling, the researchers had difficulty
constructing consistent calibration curves for both lab and full scale studies. A simple self
referencing 2-point calibration was used, assuming that the maximum detected fluorescence was
representative of the maximum concentration. Since some of the RTD characteristics predicted by
the lithium curves were not reflected in the Rhodamine curves, it appears that there may be
differences in the way that the tracers reflect mixing behavior in the IBR content matrix.

5) The lithium study implies a dead volume at all temperature levels, with dead volume being
inversely proportional with gas production as might be expected. This behavior was not
concluded by the Rhodamine studies.

6) All the studies clearly show that while approaching the ideal of a CSTR on average, mixing in the
IBR is subject to events that may impact effluent quality by providing varying degrees of mixing
at apparently random intervals. There are irregularities in the flow and mixing patterns even under
the laboratory conditions used in this work that will probably be exacerbated in real world
applications where reactors will be subjected to more complex substrates and less environmental
control. These irregularities are thought to be inherent to the low-energy input reactor design.

7) The IBR is a retained biomass reactor, and the assumption that HRT = SRT that generally follows
CSTR designation does not apply. SRT must be calculated independently using established
methods.

**Induced Bed Reactor Performance**

1) Three Induced Bed Reactor anaerobic digesters were operated successfully for 194 days at three
temperature levels (35°, 45°, and 55°C) on a completely dissolved substrate feed at maximum
OLR of 8.4 g COD-L⁻¹-d⁻¹ with a minimum HRT of 3.8 days. The most robust system, as
indicated by its ability to adapt to new loading rates with a minimum of disruption to effluent and
biogas quality was the 35°C reactor.
2) All three reactors developed beds of granulated, stratified on the bottom of the reactors. The granules were interspersed with thin rod-shaped structures that had the appearance of clipped hairs 1-3 mm in length. Bed formation demonstrated that the reactor feed systems, substrate, loading rates, and flow rates utilized for the experiments were appropriate for developing a dense granular sludge.

3) Settled sludge volumes (SSV) as determined using Imhoff cones were consistent with observations made through reactor walls; SSV for the 35°C reactor corresponded to a sludge bed occupying approximately half of the reactor volume; for the 45°C reactor, the sludge bed occupied approximately 25% of the reactor volume. The 55°C sludge bed SSV corresponded to approximately 12% of the reactor volume.

4) Flocculent sludge was also observed to accumulate in the upper chamber of the reactors above the septum. Spikes in effluent suspended solids and COD would occur periodically when these flocculated sludge beds would apparently reach a critical volume and mix with the effluent impacting both effluent quality and consistency.

5) Calculated SRTs for the reactors of 254 (35°C), 194 (45°C), and 102 (55°C) days were characteristic of retained biomass reactors and promising indicators of the suitability of the reactor design for treating high strength wastes. For purposes of design, in the absence of more permissive data the 254 day SRT observed in the 35°C reactor may be conservatively assumed to be indicative of a practical upper bound for SRT and HRT – SRT may be assumed to be the lower bound for the system SRT.

6) The IBR appears to rely on gas induced mixing to maintain contact between the sludge and the substrate in the liquid volume. Previous studies have shown that there is sufficient mixing due to heat flux through the reactor to maintain the liquid volume in a nearly completely mixed state, but the steep decline in biogas production on feed shutdown is indicative of the effective cessation of substrate reduction and the importance of contact between substrate and biomass to substrate reduction. An effective means of enhancing reactor performance might be to recirculate digester
gas through the sludge bed to enhance mixing of settled sludge and the liquid fraction of the reactor contents.

7) COD removal capacity is comparable to other high rate reactors at steady in the 35°C and 45°C reactors with observed efficiencies ranging from 73-90%. COD removal in the 55°C reactor was much lower than anticipated. The 45°C reactor was the most effective in terms of substrate reduction at the highest loading rate, but the reactor was unstable demonstrating a gradual decline in system pH necessitating a feed shutdown to permit the system to recover. Effluent COD concentrations show a significant difference in consistency as measured by the standard deviation of the distribution with the lower temperature reactors being superior in terms of producing a consistent effluent.

8) Methane production for the 35° and 45°C reactors, as calculated from the mass balance, was within the range of values expected for anaerobic digestion (57-69% of biogas by volume). The 55°C reactor methane production was significantly lower, possibly due to the effects of toxicity inhibition.

9) pH was generally stable, indicating sufficient alkalinity in the substrate to buffer any organic acid loading encountered. It was consistently observed at the high end of the 6.5-8.2 range presented by Speece (1996) as optimal for anaerobic digestion.

10) Reactor performance in terms of minimizing effluent COD decreases as loading rates are increased, but in terms of mass removal per unit volume, the highest loading rate is the most effective; the faster the substrate is run through the system, the more COD is converted even though effluent quality is worse. This has implications for designers when balancing energy production vs effluent quality.
RECOMMENDATIONS FOR FUTURE RESEARCH

1) Develop formal methodologies to accounting for hydraulic considerations in the upper volume including fluid viscosity, upflow velocity, and particle settling characteristics. These factors should be included as formal design criteria if the immobilized biomass/high SRT assumption is going to be applied, or a conservative assumption of HRT = SRT must be made to ensure that reactor performance is not impacted.

2) Develop formal methodologies for fractionation of high volatile solids substrates where a significant portion of those solids may be refractory under normal batch digestion conditions.

3) Repeat high temperature studies with experimental designs focused on developing or identifying stable digestion cultures at thermophilic temperatures. The data appears to be promising for high-throughput, high temperature digestion provided robust cultures can be developed.

4) Examine effects of enhanced mixing in the sludge bed via gas recirculation and sparging, and in the upper chamber from mechanical mixing and/or gas sparging to enhance effluent quality and consistency. Incorporate RTD studies to determine optimum level of mixing for substrate conversion and increased reactor volume utilization via elimination of observed dead zones.

5) Use of Tedlar bags for gas sampling was problematic. Tedlar bags are widely used to capture and preserve biogas samples, but they are also relatively permeable to O₂, N₂, and CO₂ three components of biogas that have relatively large concentration gradients relative to normal atmospheric gasses. This problem is not widely known and there is no literature addressing the issue. A protocol urgently needs to be developed for biogas sampling and preservation, and an appropriate study on the impact of atmospheric diffusion on the quality of anaerobic biogas samples in Tedlar bags should be undertaken.

6) Complete 16s rDNA genomic modeling of existing preserved sludge cultures from 35°, 45° and 55°C experiments discussed here. Traditionally, microbes are identified by isolating individual cultures and examining their physiological, biochemical, and morphological characteristics. More recently, a number of molecular techniques have been developed for the qualitative and quantitative analysis of microbial communities. Among them, 16S rDNA-based methods have
been applied extensively for the study of microbial diversity. This method is more reliable than microbial identification using traditional methods. Therefore, further research is proposed to investigate the phylogenetic diversity of these communities using the 16S rDNA-based technique.

7) Develop method for representative in-situ sampling of VSS to permit direct calculation of SRT and associated parameters for retained biomass reactors.

8) Investigate effects of different reactor configurations to determine whether the same mixing and conversion performance parameters observed in these studies can be maintained or improved while maximizing surface area to volume ratios to improve system economics.

9) Investigate alternative reactor heating strategies including jacketed and insulated full scale reactors.

10) Investigate feasibility of digestion of recalcitrant substrates like algae in single and multi stage processes.

11) Rhodamine WT should be studied further to determine whether it is generally suitable for anaerobic digestion tracer studies, and if so, under what conditions. Given the difficulties experienced in calibrating the tracer for this work, it may not be appropriate for anaerobic digestion studies if mass balance closure is required. This issue is not explicitly addressed in the literature.
APPENDIX A

BIOWASTE AND BIOENERGY

AN ESSAY ON THE NECESSITY OF STEWARDSHIP
BIOWASTE AND BIOENERGY:
AN ESSAY ON THE NECESSITY OF STEWARDSHIP

"...the power of population is indefinitely greater than the power in the earth to produce subsistence for man.

...A slight acquaintance with numbers will shew the immensity of the first power in comparison of the second."

--Thomas Malthus, The Principle of Population, 1798

In his classical response to the utopians of his time, Thomas Malthus pointed out the mathematical impossibility of matching an exponential consumptive growth rate with an arithmetic increase in substrate availability. Malthus did not anticipate John Snow's work on cholera, or other advances in public health. He couldn't have foreseen the mechanization of agriculture, the industrial revolution, or the green revolution. Rate of growth of agricultural production has not, to date, been strictly arithmetic. That is not to say, however, that his conclusions were incorrect.

He may have been wrong about the precise mechanism of limitation, but the laws of thermodynamics require that any system with boundaries has, by definition, limits. As such, the earth has a finite capacity to support human life. As we continue to advance our understanding of natural systems, we enhance our ability to manipulate our environment and increase our footprint as a species, but we also begin to see that there may be inhibitory limits other than substrate production that must also be considered as concentration of waste products impacts our species.

Societal recognition of this principle has driven the creation of laws addressing environmental preservation and agencies to enforce those laws. This has led to increased focus on and understanding of environmental issues in the scientific community. A huge amount of effort is being expended, both politically and scientifically in an attempt to understand the interplay between environment, energy, and populations, and how we can successfully address the problems that are presented to us by our desired patterns of consumption and the outcomes of those patterns.
Fundamental to this work is an understanding of energy on a macro-level as it impacts human populations. For all practical purposes, all energy is solar in origin. Energy from the sun strikes the earth at a rate of 178,000 TW (Davis, 1990) where it is either captured (as heat via radiation or chemically via photosynthesis), or radiated back into space. It is estimated that all the biomass on the planet utilizes 100 TW (Nealson and Conrad, 1999). This energy finds its way into the biosphere via photosynthesis for direct conversion to energy to drive life processes, and to store energy in chemical bonds. It is true that there are life forms that extract carbon and energy from other sources, but the vast majority of life on the earth receives its energy either directly from photosynthesis or by consuming something that does.

Human populations consume the energy captured thusly as food at a rate of 0.96 TW as calculated from calorie consumption (WHO, 2003). Of this, approximately 77% is derived directly from plant matter, while 23% is derived from consumption of other heterotrophs (ibid). Given that it requires from ten to twenty times the land area to generate as much energy as animal biomass as vegetable biomass (Spedding, 1990), human food consumption of heterotrophs may be responsible for an additional 2-4 TW of direct solar energy for a total consumption rate of 2.7-4.7 TW. Assuming that all this is correct, the human race consumes as food approximately 3-5% of the energy captured phototrophically by the biosphere.

That is not, however, the only energy that we use. Every day a fraction of the energy captured by biomass makes its way into the relatively stable but high-energy carbon molecules that constitute fossil fuels. Consumption of fossil energy is approximately 13.6 TW (USDOE, 2006). Estimates of world fossil energy reserves are $5.5 \times 10^{10}$ TJ (USGS, 2000; USDOE, 2006), but assuming a billion years for accumulation, that yields an accumulation rate of $0.000 \ 0017$ TW, or 1.7 MW. This is roughly $1/8,000,000$ of the current rate of consumption, and $1/60,000,000$ of the rate at which energy is made available in the environment. Even if recoverable reserve estimates are off by an order of magnitude, fossil fuels are clearly not a reliable steady state resource. On top of the inherent inefficiency, there is a consensus in the scientific community that carbon released by fossil fuels has a negative impact as it accumulates as CO$_2$ in the environment.

Nealson and Conrad (1999), in their definition of life, state that it “…must develop some means for escaping from its own metabolic end products.” Malthus was incorrect about growth of agronomic rates in
1797, but his premise may hold true with regard to other limiting factors. Humans, while probably not evolving much biologically over the past two centuries, have certainly evolved technologically. It remains to be seen whether the biological organism is sufficiently robust to continue to expand in the environment created by the technological symbiote. In the meantime, however, it behooves us to learn to adapt to the sustainable resources that we can access.

If we were to tap into the 178,000 TW of energy that strike the planet on any given day and in any meaningful way, the carbon question could become a thing of the past, and we could proceed to test Mr. Malthus in new ways. If we could even get to a portion of the 100 TW already being absorbed into the biosphere, we could make some progress on the question of carbon. Much effort is being placed in these areas. In the meantime, however, we have existing “metabolic end products” to escape from as we discharge high strength wastes from our biological and technological metabolisms into the environment (Chen, 2003). Another way of looking at “high strength” however is “high energy”. By approaching the problem from the standpoint of energy, we can address two problems with one solution.

Prior to the advent of the germ theory of disease and the growth of public health awareness, it was assumed (if anyone thought about it at all) that the environment had sufficient capacity to stabilize wastes. There are at least three reasons to control the degradation of high strength organic wastes. First and most directly, pathogen and thus disease control is possible if wastes are oxidized to a low energy state in a controlled environment and kept separate from substrate sources. Second, controlled treatment permits predictive environmental conditions by ensuring that shock or steady state loadings to receiving environments do not exceed a level that provides acceptable environmental results (Janzen, 1999). Finally, there is a significant amount of energy available in organic waste discharges; it is estimated that if 50% of the manure in the United States were processed into fuel it would displace 5% of US coal consumption (Karim et al., 2007). High energy substrates will be utilized by organisms, whether in the environment or in a controlled process. If those energies are controlled, they can be extracted and used to displace fossil fuels, resulting in a net zero carbon emission.

It behooves us to develop ways to do a better job of maintaining a stable and beneficial environment capable of steady-state agronomic and reproductive rates. To do this we must become better stewards of
the energy we capture, and learn to cycle it as efficiently and holistically as our understanding will permit. Statistical data shows a strong correlation between the increased global standard of living and increased consumption of animal products with their inherent inefficiencies as energy sources (UNFAO, 2006). If we permit food preferences to shift to less efficient sources, then we also need to do our best to limit the impact of those preferences on our constructed and natural energy economies.

We've been successful in extending the horizons of Malthus' prediction by focusing on substrate (food) development via land reform, various agro/industrial advancements, the green revolution, and GMO crop development. Over the past 40 years, however, we've also become increasingly aware of the importance of moving away from our own inhibitory wastes, hence the legal and social framework of the environmental Green and Blue movements. While we have made progress in waste stabilization, we have not paid sufficient attention to not only the detrimental impacts (viz. pathogen control, directly observable environmental costs) of the waste if left untreated, but also the benefits of cycling carbon and energy, and the less obvious impact that those have on our biosphere.

REFERENCES


APPENDIX B

RHEOLOGICAL PROPERTIES OF AGRICULTURAL SLURRIES INCLUDING DAIRY MANURES:

CONSIDERATIONS FOR FLUID HANDLING DESIGN
RHEOLOGICAL PROPERTIES OF AGRICULTURAL SLURRIES INCLUDING DAIRY MANURES:

CONSIDERATIONS FOR FLUID HANDLING DESIGN

INDUCED BLANKET REACTOR

WHITE PAPER # 1

J. SHAUN DUSTIN, M.S., P.E.

INTRODUCTION

In design of an anaerobic digestion system for high solids content substrates, consideration must be given to rheological properties of the feed, reactor contents, and digestate. Viscosity of fluids impacts mixing of the substances to be reacted, pump type and size, pipe sizing, and unit process design for intermediate processes such as heat exchange.

Dairy manures and other organic slurries can be classified in terms of rheological properties as non-Newtonian fluids. This means that the relationship between shear stress and shear rate is not constant. Their viscosity increases as a function of solids content, and decreases as a function of temperature. Viscosity influences pressure drop and flow regime for fluid, impacting pumping, transport, up- and downstream processes, and mixing in a reactor.

FLUID CLASSIFICATION

El-Mashad (2005) demonstrated that the relationship between shear stress and shear rate is only linear at higher rates of shear (more than 50 s\(^{-1}\)), but even then, it varies as a function of temperature. The higher the temperature, the lower the apparent viscosity, with the temperature effect being described by an Arrhenius-type model. This demonstrates the non-Newtonian nature of the fluid. El-Mashad (2005) and others (Chen and Hashimoto, 1976; Hashimoto and Chen, 1976; Schofield, 1984; Chen, 1986; Moeller and Torres, 1997) confirm this result for other organic slurries including manures, wastewater sludges, and other organic sludges.
El-Mashad postulates that this generalized behavior may be a function of the presence of relatively large particles in the fluid. Hashimoto and Chen (1976) found that for their work with livestock wastes, larger floc particle sizes translated to increased apparent viscosities for fluids having the same solids content.

El-Mashad also showed that dairy manures behaved as real plastic fluids, and compared his results to other waste slurries to show that they fit the same power law model. Beef manure (Hashimoto and Chen, 1976; Chen, 1986), anaerobically digested municipal sludge (Moeller and Torres, 1997), and Pekmez (a grape product produced in Turkey) (Kaya and Belibagli, 2002) were all considered, and shown to fit the same general equation:

\[(0.1) \quad \tau = k \lambda^n\]

\[(0.2) \quad k = -0.3T + 29\]

Where \(\tau\), shear stress (Pa); \(\lambda\), rate of shear (s\(^{-1}\)); \(k\), consistency coefficient (Pa s\(^n\)); \(n\), flow behavior index and \(T\)=temperature (°C).

Solving for \(n\) provides an indication of the sensitivity of viscosity on rate of shear, although it does not directly produce the viscosity. El-Mashad calculated values ranging from 0.211-0.342 for dairy manure. A value of 1 indicates Newtonian behavior. These slurries are non-Newtonian with distinct real plastic behavior.

**DESIGN CONSIDERATIONS**

Schofield’s review of agricultural slurry properties (1984) is substantiated by El-Mashad’s conclusions, but cautions workers to be aware of the wide range of solids contents, solids types, and temperature ranges encountered in agricultural slurries, and to account for them in design. He points out the disparity between reported values for viscosity as a function of total solids (TS) (up to two orders of magnitude for the same reported TS value), temperature, and pH. These differences make generalization of fluid properties difficult, and may invalidate some “rule of thumb” design procedures in common practice (viz. “viscosity of water plus 10%”). Researchers have observed pressure drops of 2x to 12x those predicted when using
the viscosity of water as the design criteria under laminar flow conditions. Head losses can be minimized by maintaining turbulent rather than laminar flow.

Schofield also cautions researchers to be aware of the limitations of rotary and small scale capillary viscometers in the analysis of high solids content slurries as used by El-Mashad and Chen, given the large particle sizes and high solids contents of the fluids under consideration.

For maintenance of solids any fluid transport or reaction system, fluids may be classified as either microfluids, meaning that they are free to mix on a molecular level, or macrofluids, meaning that they have a tendency to clump or aggregate and mix as distinct particles with surface areas small relative to their volumes (Levenspiel, 1999). In-tank mixing design should take into account the viscosity of the fluids involved. Four distinct phases are present:

1) **Feed.** Feed may be classified as a macrofluid or microfluid depending on its characteristics. It is assumed that for the high solids substrates that are the focus of this presentation, the feed is a non-Newtonian real plastic fluid. While this does not require that the fluid behave as a macrofluid, it is increasingly likely as the apparent viscosity increases, and should probably be treated as such.

2) **Retained biomass.** In an IBR, a sludge bed develops in the bottom of the reactor. This sludge bed is comprised of flocs or granules of agglomerated biomass. The feed filters through this active biomass, which catabolizes the feed first to organic acids and methane precursors, then to methane and carbon dioxide biogas. The retained biomass particles are by definition a macrofluid

3) **Biogas.** Biogas is considered a macrofluid, although it is chemically functionally inert in the reactor. It is generated throughout the reactor volume but primarily in the sludge bed as a function of the reactions between the feed and the active biomass. It accumulates in the sludge bed, and boils off, rising through the water column and inducing shear and mixing as it transits the reactor volume. Mixing in the IBR is derived primarily from this action coupled with the convective mixing induced by the thermal gradients in the reactor.

4) **Digestate (effluent).** The digestate is the fluid between the sludge bed and the reactor outlet. The digestate may be a macrofluid or a microfluid depending on the viscosity change brought about by the digestion process.
RECOMMENDATIONS

Designers need to be aware that high-solids organic slurries will not behave like water, and design accordingly. For optimal design, an understanding of the rheological characteristics of the fluid to be pumped should be developed in accordance with the methods used by El-Mashad, but using a large scale tube viscometer similar to that described by Cumby (1980).

In the absence of such data, designers should recognize the non-Newtonian real plastic character of the fluid and select equipment accordingly. The values proposed by El-Mashad may be acceptable, provided that the temperature correlation holds true for the full range of anticipated temperatures since pump and conveyance design will be controlled by maximum viscosity which will occur at minimum fluid temperature and maximum solids content. It should also be recognized that viscosity is also a function of shear rate, with apparent viscosity decreasing as shear rate increases.

For most cases, the pump will be moving the fluid from a low temperature reservoir to a higher temperature steady state, and the pumps and conveyances must be designed for the low temperature condition. The pump should be designed to move the fluid efficiently and reliably at the coldest anticipated operating temperature with the highest desired solids content. This indicates a design temperature of 2°C and a solids content of 10% TS for an IBR.

For transport a Reynold's number (R) in excess of 4300 should be maintained to ensure turbulent flow in the pipes. This permits minimization of pressure drop and prevents sedimentation in the pipe by maintaining solids in suspension (Chen and Hashimoto, 1976).

Digesters generally rely on a constant temperature feed. Process heat is supplied via a liquid/liquid heat exchanger. The heat exchanger passages for the cold substrate fluid should be sized like the pipes to maintain turbulent flow and to pass the largest solids anticipated in the lines at the lowest temperature anticipated. Even if the warmer fluid will permit reducing pipe size through the heat exchanger, this is not recommended.

In an IBR, if the viscosity of the feed and sludge bed are such that mixing is minimized or the feed fluid properties are such that the sludge bed cannot be maintained as a distinct zone in the reactor, reactor efficiency will suffer.
SUMMARY

In summary, it is difficult to predict apparent viscosity of real slurry systems for mixing and feed design into anaerobic digesters without site-specific fluid characterization. In the absence of definitive data to the contrary, designers should therefore assume real plastic non-Newtonian behavior for organic slurry fluids and design digestion vessels and supporting systems accordingly with appropriate factors of safety and equipment selections in order to provide consistent operation (Schofield, 1984). El-Mashad provides values for apparent viscosity of 10% solids dairy manures at temperatures ranging from 30-60°C, and a methodology for deriving apparent viscosity for other circumstances.

REFERENCES


APPENDIX C

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Title PhD candidate, Utah State University
7 January 2010

To Whom It May Concern:

I hereby give my permission to Jacob Shaun Dustin to include the paper titled “Introduction to the IBR”, of which I am a co-author, in his dissertation.

Please feel free to contact me if you have any questions regarding the above at 208-520-8725 or via email at here17@gmail.com.

Respectfully,

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CURRICULUM VITAE

Jacob Shaun Dustin

(January 2010)

Candidate for the Degree of Doctor of Philosophy


Major Field: Biological Engineering

EDUCATION

Ph.D. Biological Engineering. 2010. Utah State University, Logan, Utah.
M.S. Civil Engineering. 1998. Brigham Young University, Provo, Utah.
B.S. Civil Engineering. 1998. Brigham Young University, Provo, Utah.

AWARDS AND CERTIFICATIONS

Professional Engineer (CA, ID, MT, UT, WY, MN, WA, OR)
USTAR Assistantship, Utah State University (2006-2009)
USDA Natural Resources Conservation Service Technical Service Provider, waste management and energy generation (2007)
Idaho Society of Professional Engineers, SE Chapter Young Engineer of the Year (2001)
Departmental Graduate Scholarship, Brigham Young University (1996-98)
Leadership Academy Scholarship, Brigham Young University (1990-91)
Eagle Scout (1986)
ACADEMIC EXPERIENCE

Utah State University Research Foundation Energy Dynamics Laboratory. 2009-10.

Responsible for design and development of pilot scale (2000,000 gpd) waste to energy plant (RENEWER) under an ARRA grant from the USEPA, administered by UTDEQ. Work required design, construction operation, and analysis of results for integrated process train including algal growth raceways, dissolved air floatation, and anaerobic digestion. Supervised graduate and undergraduate research, and designed and developed a mobile analytical laboratory.

Utah State University. 2007-08.

BIE 5610 “Food Engineering”

Developed syllabus and lectures for process thermodynamics portion of course.

Presented 20 lectures, developed exams, and administered grades for this work.

NFS 4440 “Fundamentals of Food Engineering”

Developed syllabus and lectures for process thermodynamics portion of course.

Presented 20 lectures, developed exams, and administered all grades for this work. Led field trips to food processing facilities.

Brigham Young University. 1995-98.

Statics, Surveying, Pavement Design.

Office hours. Graded homework. Assisted with Labs.

Structures Laboratory

Laboratory Technician; Collaborated with faculty to implement experimental designs for structural and geotechnical experiments. Work included construction, instrumentation, and controls. Aided staff in maintaining and operating all department lab facilities.
PUBLICATIONS

Refereed Journal Articles

Dustin, J. S., Hansen, C.L., Thermophilic Operation of the Induced Blanket Reactor Anaerobic Digester, Transactions of the ASABE, in review.


Conference Proceedings


Podium Presentations

Thermophilic Operation of the Induced Blanket Reactor Anaerobic Digester, ASABE Annual International Meeting, Reno, NV June 2009


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Seismic Design Report, Miller Children's Hospital, Long Beach, CA 2007

Hardy Frame Moment Frame Catalog, Hardy Frame Inc., 2006

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Seismic Design Report, St. Joseph Medical Office Bldg, Orange, CA 2006

Seismic Design Report, St. Joseph Cancer Center, Orange, CA 2006

Seismic Design Report, United States Federal Courthouse, Los Angeles, CA 2006

Seismic Design Report, Yale New Haven Hospital, New Haven, CT 2005

Seismic Design and Construction Oversight, Magic Technologies Clean Room, Milpitas CA, 2005

Seismic Design and Construction Oversight, AMC Phase 1B Century City, CA 2005

Seismic Design, Salt Palace Phase 3 Design-Build, Salt Lake City, UT 2005

Seismic Design and Construction Oversight, Los Robles Hospital, Thousand Oaks, CA 2005

Seismic Design and Construction Oversight, African American Unity Center Seismic Retrofit, Los Angeles CA 2004

Seismic Design and Construction Oversight, Overlake Hospital Tower, Bellevue, WA 2004

Seismic Design and Construction Oversight, Mission Hospital Medical Office Bldg, Mission Viejo, CA 2004
Seismic Design and Construction Oversight, Alexander Residence, Calabasas, CA 2004

Seismic Design and Construction Oversight, Dougherty Residence, Long Beach CA 2004

Seismic Design and Construction Oversight, Salt Lake Tribune and Deseret News Press Hall Production Facility, Salt Lake City, CA 2004

Seismic Design and Construction Oversight, Ebell Theater Loft Conversion, Long Beach, CA 2004

Seismic Design and Construction Oversight, Galpin Auto Sport, North Hill CA 2004

Seismic Design and Construction Oversight, Disneyland Space Mountain Retrofit of Load/Unload Structure, Anaheim, CA 2004


SidePlate Special Moment Frame Connection Design Basis for OSHPD, 2004

LEADERSHIP

City Councilman, Nibley UT 2008-Present

Planning and Zoning Commissioner, Nibley, UT 2007

President, SE Idaho Chapter, Idaho Society of Professional Engineers 2001-02

Vice President, ASCE student chapter, Brigham Young University

Upper Snake River Resource Advisory Council, Bureau of Land Management 2000-03


Boy Scouts of America, Scoutmaster, 1998-Present

Student Body President, Woodbridge High School, RAF Woodbridge, UK 1989-90
CONSULTING EXPERIENCE

Principal, Dustin Engineers. 2006-2009

- Design and installation oversight of federally funded anaerobic digestion system for waste treatment and energy production at the Huls Dairy in Hamilton, MT.
- Process design for integration of existing wastewater treatment plant with algae to energy project for USU Energy Dynamics Lab and City of Logan, UT.
- Engineering support for development of the Shoshone Renaissance Geothermal Field, Honeyville, UT.
- Environmental assessments for the Northwest Band of the Shoshone Nation and City of Logan, UT

Vice President, SidePlate Systems. 2003-2006

- Address concerns of owners, investors, regulators and constructors to ensure smooth implementation of new technologies.
- Responsible charge of seismic structural engineering projects including courthouses, schools, and hospitals throughout the US. Clients included GSA, Walt Disney Imagineering, and Kaiser.
- Responsible for new IP development and deployment

Civil Engineer, Intrepid Engineering Services. 2000-2003

- Developed and implemented program for field investigation of 900+ beneficial use water rights for the Idaho Department of Water Resources Snake River Basin Adjudication.
- Feasibility assessment of multiple energy projects in Southern Idaho including anaerobic digestion, hydropower, and ethanol.
- Responsible charge for Civil portion of in-situ groundwater remediation and site remediation for cleanup of 3-Mile Island waste at INL.
Civil Engineer, Schiess & Associates. 1998-2000

- Project engineer for multiple USDA-RD funded water and wastewater development projects in southeastern Idaho.

- Field engineer for wastewater treatment plant upgrades in Pocatello, Blackfoot, and Fort Hall Idaho.

- Well design and development.

- Principal author, Pocatello Stormwater Master Plan, Blackfoot Water System Master Plan

LANGUAGES

English—Native language

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MEMBERSHIPS

American Society of Agricultural and Biological Engineers

American Society of Civil Engineers

National Society of Professional Engineers

Institute of Biological Engineers

American Water Works Association

Water Environment Federation

Boy Scouts of America