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## **Two Novel Floor-Scale Anaerobic Digester Systems for Processing Food Waste**

**Part 1: Multi-Reactor, Liquid Recycle System for  
High Solids**

**Part 2: Moderate Solids Digester System**

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# **Two Novel Floor-Scale Anaerobic Digester Systems for Processing Food Waste**

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## **Part 1: Multi-Reactor, Liquid Recycle System for High Solids**

## **Part 2: Moderate Solids System**

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# Table of Contents

	<u>Page</u>
List of Figures and Tables.....	iv
Figures.....	iv
Tables.....	vi
Acknowledgements.....	vii
From Rotting Food Scraps to Sustainable Energy, Fertilizer and Compost: How Research is Getting Us There.....	viii
Executive Summary .....	xiii
Problem-solving approach: multi-reactor, liquid recycle system for high solids .....	xiv
Pilot project approach: moderate solids digester system .....	xvii
Part I: Multi-Reactor Liquid Recycle System for High Solids .....	1
Section 1: Concerns with composting food scraps and fresh grass .....	5
Background.....	5
Compost of food scraps and fresh green waste .....	6
Incorporation of anaerobic digestion with compost .....	12
Section 2: Anaerobic digestion of food scraps and green waste.....	14
Background.....	14
Anaerobic digestion economics.....	15
Technologies used for anaerobic digestion of food waste and green waste.....	17
Types of non-slurry systems.....	19
Conclusion.....	23
Section 3: Floor scale evaluation of multi-reactor, liquid recycle high solids anaerobic digestion system.....	25
Background.....	25
New leaching bed reactor concept.....	25
Theory.....	27
Goal and objectives .....	28
Section 4: Mixed food scrap and green waste characterization, specific and ultimate methane potential and nutrient recovery capability.....	30
Background.....	30
Synthesis of solid mixture .....	30
Characterization of solid mixture and other digester inputs.....	31
Biogas and methane productivity calculations .....	32
Section 5: Experimental methods .....	34
Experimental system .....	34
Experimental design .....	36
Analytical methods .....	36
Section 6: Batch digestion with only internal leaching bed reactor recycle .....	38
Background.....	38
Results and discussion.....	38
Conclusion.....	43

Section 7: Batch digestion using both internal leaching bed reactors and external upflow anaerobic sludge blanket feed .....	44
Background.....	44
Results and discussion.....	45
Conclusion.....	52
Section 8: Batch digestion using phased internal leaching bed reactors and external upflow anaerobic sludge blanket recycle .....	53
Background.....	53
Results and discussion.....	53
Conclusion.....	58
Section 9: Full system performance.....	60
Background.....	60
Results and discussion .....	60
Section 10: Scale-up and conclusions.....	63
Scale-up .....	63
Project conclusions .....	65
<b>Part II: Moderate Solids System .....</b>	<b>67</b>
<b>Section 1: Introduction.....</b>	<b>70</b>
<b>Section 2: Project objectives .....</b>	<b>71</b>
<b>Section 3: Project approach .....</b>	<b>72</b>
The design concept of the floor scale system, and preliminary optimization for the floor-scale system using laboratory-scale experimentation .....	72
Testing of floor scale moderate solids dual reactor system.....	74
Testing of intermittent mechanical mixing.....	75
Development of the process software model .....	76
Nutrient recovery.....	77
<b>Section 4: Project results.....</b>	<b>81</b>
Objective 1: Floor-scale design and build-out of the moderate solids system, and preliminary optimization for the floor-scale system using laboratory-scale experimentation .....	81
Objective 2: Floor-scale testing of the moderate solids system .....	85
Objective 3: Effect of mechanical mixing in the laboratory-scale experiments	90
Objective 4: Development of a mathematical model for the moderate solids anaerobic digester system and operational insights for scale-up .....	92
Objective 5: Nutrient recovery after floor-scale testing of moderate solids system .....	97
Objective 6: Disseminate scientific publications developed as a result of this project .....	98
<b>Section 5: Conclusions and recommendations for pilot scale design .....</b>	<b>99</b>
Additional recommendations for pilot-scale design.....	100
Public benefits to Washington.....	102
<b>References .....</b>	<b>103</b>
<b>Appendices.....</b>	<b>113</b>
Appendix A. Floor scale moderate dual reactor system drawing .....	113

Appendix B. Numerical simulation of mechanical mixing in high solid anaerobic digester .....	116
Appendix C. Lipid production with dark fermentation hydrogen production effluent using oleaginous yeast <i>Cryptococcus curvatus</i> .....	117
Appendix D. Multiphase modeling of mixing effectiveness in anaerobic sequencing batch reactor (ASBR) .....	118
Appendix E. Evaluating rate-limiting step in anaerobic digestion of flushed dairy manure and the effect of microbial community ratio on its kinetics .....	119
Appendix F. Optimize butyric acid and bio-hydrogen production from food waste using untreated UASB effluent as seed .....	120

# List of Figures and Tables

## Figures

	<u>Page</u>
Figure 1: Roosevelt Regional landfill Washington State (photo: WA Ecology, 2010b).....	5
Figure 2: LCA for various food scrap treatment options (Kim and Kim, 2010) .....	6
Figure 3: Compost facility schematic (Biocycle, 2008) .....	8
Figure 4: Evolution of ammonia-N as percent of initial total N (MXP: mixed; Seed: aged compost used as initial seed; FW: food waste with no-seed: FW <sub>ns</sub> : food waste without seed; YW: yard waste with seed; YW <sub>ns</sub> : yard waste without seed; YWh: household yard waste; EPA, 2003) .....	10
Figure 5: Anaerobic digestion process (Bitton, 1994) .....	14
Figure 6: Capital cost by design capacity and operating costs by flow throughput for commercial OFMSW digesters as reported in Hartmann and Ahring (2006). ....	16
Figure 7: Summary of various OFMSW AD technology approaches .....	18
Figure 8: Trends in Slurry (Wet) and Non-Slurry (Dry) AD of OFMSW .....	19
Figure 9: Varying digester designs in single phase, single reactor approach (A is Dranco; B is Kompogas, C is Valorga, from Lissens et al., 2001) .....	20
Figure 10: Commercial application of single-phase and two-phase systems by year as against total installed capacity (Hartmann and Ahring, 2006).....	21
Figure 11: SEBAC system as example of sequential use of single reactor, phased LBR (Chynoweth et al., 1992).....	22
Figure 12: Schematic, flow pattern and design approach of new LBR system for treatment of food scrap and green waste composed OFMSW within a compost yard scenario .....	26
Figure 13: Overview of methane specific biogas productivities at commercial facilities, plotted against OLR, waste mixture, and volumetric biogas production (Hartmann and Ahring 2006) ..	30
Figure 14: Example of one of the solid waste mixture buckets collected, stored and utilized during project research.....	31
Figure 15: HSAD Experimental system .....	34
Figure 16: LBR reactor bucket design and operation .....	35
Figure 17: pH against reaction time for non-acclimated and acclimated batch runs .....	39
Figure 18: VFA against reaction time for non-acclimated and acclimated batch runs.....	39
Figure 19: VFA against reaction time with change in percent acetate for non-acclimated batch	40
Figure 20: VFA against reaction time with change in percent acetate for acclimated batch.....	40
Figure 21: Biogas accumulation against reaction time for non-acclimate and acclimated batch runs.....	41
Figure 22: Biogas accumulation against reaction time with changes in biogas constituency for non-acclimated batch run .....	41
Figure 23: Biogas accumulation against reaction time with changes in biogas constituency for acclimated batch run .....	42
Figure 24: pH against reaction time for 1x, 2x, and 4x feed rates each at 4 g/L alkalinity .....	45
Figure 25: VFA against reaction time for 1x, 2x, and 4x feed rates each at 4 g/L alkalinity.....	46
Figure 26: Biogas accumulation against reaction time 1x, 2x and 4x feed rate and 4 g/L alkalinity .....	47

Figure 27: Biogas accumulation against reaction time with changes in biogas constituency for 2x feed rate and 4 g/L alkalinity .....	48
Figure 28: VFA against reaction time with change in percent acetate for 2x feed rate at 4 g/L alkalinity .....	49
Figure 29: pH against reaction time for 4 and 8 g/L alkalinity, each at 1x feed rate.....	50
Figure 30: pH against reaction time for 4 and 8 g/L alkalinity, each at 4x feed rate.....	50
Figure 31: VFA against reaction time for 4 and 8 g/L alkalinity, each at 1x feed rate .....	51
Figure 32: VFA against reaction time for 4 and 8 g/L alkalinity, each at 4x feed rate .....	51
Figure 33: pH against reaction time for full batch system using either acclimated or non-acclimated seed .....	54
Figure 34: VFA against reaction time for full batch system using either acclimate or non-acclimated seed .....	55
Figure 35: VFA against reaction time and change in percent acetate for full batch system using acclimated seed (16.7% biomass) .....	55
Figure 36: Biogas against reaction time for full batch system using either acclimated or non-acclimated seed .....	56
Figure 37: Biogas against reaction time and changes in components for full batch system using non-acclimated seed (UASB granules) .....	57
Figure 38: Biogas against reaction time and changes in components for full batch system using acclimated seed (16.7% biomass) .....	57
Figure 39: Comparison of methane percentages against time for full and internal recycle only systems .....	58
Figure 40: Scale up of single phase, dual reactor AD process for treatment of food/green waste	63
Figure 41: Flow chart of floor scale moderate dual reactor design .....	72
Figure 42: The floor scale moderate solids anaerobic digestion system .....	74
Figure 43: Process model of the floor scale moderate solids anaerobic digestion system .....	77
Figure 44: IMARNRS pilot testing facilities .....	78
Figure 45: Schematic Diagram of IMARNRS .....	78
Figure 46: The lab-scale nutrient recovery system .....	80
Figure 47: Natural separation of liquid and solid in a mixture of food waste and inocula from the effluent of UASB seed reactors (left bottle), and food waste and water (right bottle). (Left picture: after one day, Right picture: after two days).....	82
Figure 48: Change of total VFA with time at the different pH .....	83
Figure 49: Change of VFA concentrations with pH .....	83
Figure 50: Change of TS reductions with pH .....	84
Figure 51: Effect of recirculation ratio on the pH of the high solids digester.....	86
Figure 52: Daily biogas productions in the high solids digester (experimental data).....	87
Figure 53: Change of COD and VFA over time in the UASB seed reactors .....	87
Figure 54: Cumulative biogas production at a feeding rate of 10 gal/week in the UASB seed reactors .....	88
Figure 55: Cumulative biogas production at a feeding rate of 10 gal/day in the UASB seed reactors .....	88
Figure 56: Daily biogas production in the UASB seed reactors .....	89
Figure 57: Effect of mixing on VFA and pH .....	91
Figure 58: Comparison of daily biogas production between prediction and experimental data from UASB seed reactors .....	93

Figure 59: Comparison of VFA between prediction and experimental data from the UASB reactor .....	94
Figure 60: Predicted effect of recirculation rate on pH (A) 0.09 m <sup>3</sup> /day (B) 0.13 m <sup>3</sup> /day (C) 0.19 m <sup>3</sup> /day.....	95
Figure 61: Predicted effect of recirculation rate on acetic acid concentration (A) HSAD (B) HSAD bottom (C) UASB .....	96
Figure 62: Flow chart of pilot scale moderate solids anaerobic digester system .....	101

## Tables

Table 1: Characteristics of food scraps and fresh green waste .....	7
Table 2: Preferred operating conditions for compost of low pH, high VFA food scrap/fresh grasses containing bio-waste.....	12
Table 3: Influent parameters .....	32
Table 4: Expected minimum range of performance outcomes for reactor system and feed.....	33
Table 5: Experimental plan and treatments .....	36
Table 6: Mass balance of full system for LBR and external UASB components (37 days).....	60
Table 7: Nutrient flows .....	61
Table 8: Gas production performance indicators for full system by component and total (37 days) .....	62
Table 9: Observations of food waste with different total solids (TS) percentages .....	76
Table 10: Characteristics of diluted manure and kitchen waste .....	79
Table 11: Comparison of new HSAD and APS-digester in system characteristics .....	85
Table 12: Comparison of new HSAD and existing systems in productivity .....	90
Table 13: Conditions for the mathematic model for the moderate solids anaerobic digester system. ....	93
Table 14: Ammonia stripping efficiency at different TAN concentrations .....	97
Table 15: N and P of nutrient recovery for food waste from UASB effluent .....	98

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# **From Rotting Food Scraps to Sustainable Energy, Fertilizer and Compost: How Research is Getting Us There**

## **A Preface to the High Solids Anaerobic Digestion Technical Report**

In recent years, there has been a growing interest and public dialog regarding the sustainability of the food we consume. However, there has seemingly been far less interest in the sustainability of food *after* consumption – at least until managing that residual food waste becomes a problem our noses simply cannot ignore. Managing residual organics from our cities, particularly the highly putrescible food waste residuals, is a major sustainability challenge not unlike developing local and regional food sources for cities.

26-40% of all food in the U.S. is wasted<sup>1</sup> with energy embedded in that waste representing 2% of annual U.S. energy consumption<sup>2</sup>. In Washington State, 1 million tons of food is landfilled annually<sup>3</sup>, contributing to greenhouse gas (GHG) emissions as well as many other environmental concerns. For perspective, diversion and composting Washington State's annual food scraps would reduce GHG emissions by 872,695 MT CO<sub>2e</sub>, representing 1.8% of Washington State's target reduction by 2050<sup>4</sup>.

Recycling food waste can also provide economic benefits. A 2001 California study estimated that diversion and recycling of waste nearly doubled economic benefits relative to landfilling that same material<sup>5</sup>. A Seattle study found that food scrap collection and diversion programs represent a 20% reduction in cost relative to landfilling<sup>6</sup>.

Washington citizens can take pride in the fact that we, along with rest of the West Coast, lead the U.S. in implementing food waste “recycling” programs<sup>7</sup>. These programs, though, have encountered significant challenges to operationalizing sustainability goals. Cities including Seattle, Toronto, Seoul, and Stockholm have all experienced public concerns regarding odor, air emissions and public health due to the increased flow of food waste diverted to regional composting facilities. In many cases, this food waste has already begun decomposing prior to arrival at the compost facility due to storage and transportation timelines inherent in the collection process. Particularly when coupled with large inflows of grass clippings as temperatures warm in the late spring to early summer, this can dramatically increase the potential for odor generation.

Furthermore, recent studies from Swedish researchers<sup>8</sup>, the U.S. Environmental Protection Agency<sup>9</sup>, and the U.S. Compost Council<sup>10</sup> demonstrate that inclusion of partially decomposed food waste can lead to rapid pH depression, which alters the compost pile biology, resulting in conditions that actually *inhibit* standard composting processes. Operation under such a situation leads to the increased release of Volatile Organic Compounds (VOC's), methane, and ammonia – all potential sources of potent odors and other environmentally harmful gases. These studies further indicate that standard compost management strategies (increasing compost pile

temperature and air-flow) designed to reduce odor problems under normal composting conditions may, in fact, *increase* the problem due to microbial process inhibition.

Current strategies for mitigating these “new” challenges created by food waste recycling have focused on upgrading investments in process and odor control technology to improve material handling across the whole compost facility. Specific actions include determining new optimal compost pile sizes, effective C/N ratios, moisture control mechanisms, and phasing of temperature and aeration treatments. The City of Toronto recently required compost facilities within the region to make extensive and costly infrastructure upgrades (more indoor, air-controlled processing, upgraded ammonia recovery and bio-filters, improved exhaust stacks) while simultaneously and significantly reducing the overall material flow into existing compost facilities<sup>11</sup>.

Because these types of changes create negative impacts on the financial sustainability of compost facilities, many organic waste managers are looking for alternative processing technologies better suited for food waste. In Seoul and Stockholm, national research funding has been designated for technology development for new food waste processing strategies, such as anaerobic digestion and conversion of leachates and solids to animal feed.

Anaerobic digestion – decomposition of organic wastes in the absence of oxygen - offers strong potential to integrate with existing composting technology to reduce GHG emissions, control odor and air emissions, and reduce human health concerns while harnessing food wastes for energy recovery. Combining anaerobic digestion and composting enables organic waste processors to better balance material flows, utilizing the most appropriate technology for primary treatment of each waste type while creating synergy for the whole process and final product quality<sup>12</sup>.

For instance, wet organic materials such as food waste can be immediately directed to the anaerobic digester in a contained receiving building. The highly volatile (odor, ammonia and GHG generating) material is stabilized through anaerobic digestion and recovered as renewable energy in the form of methane, and liquid fertilizers are separated. The stabilized outputs from the digester can then be added to dry, woody and fibrous green waste materials in a traditional composting process for final processing without the down-side odor problems associated with composting highly volatile acidic food wastes. A 2001 study comparing a combined anaerobic digestion and compost operation to composting-alone showed a *sevenfold decrease* in releases of ammonia and VOC<sup>13</sup>. A Swedish study on GHG emissions showed that the lowest emissions resulted from combined digestion and composting relative to composting alone or land-filling<sup>14</sup>.

Unfortunately, there are still obstacles to achieving this powerful integration of anaerobic digestion and composting in commercial organic recycling facilities in the U.S., not the least of which is the financial commitment. Therefore, no combined systems have yet been installed in an organics recycling facility in the U.S. Compost facilities themselves are both capital and operating cost intensive, but the inclusion of commercially available anaerobic digestion systems to process food waste can double these costs. For perspective, a typical 100 MT/day anaerobic digestion unit appropriately sized for West Coast cities would cost \$10-15 million to construct

and around \$0.5M annually to operate<sup>15</sup>. European digesters have been economically viable due to much higher electrical power rates than are generally available in Western North America.

There are also processing challenges that need to be overcome for successful digestion of energy-rich food waste in commercially available technology. Food waste, while more readily degradable than manures (traditional digestion feedstock), also are much less stable in the digestion process. This can lead to rapid biological reaction, souring, and inhibition of the digestion process<sup>16</sup>. Therefore, process control mechanisms must be built into the digestion technology to create a stable process for methane production. Contaminants (i.e. plastics, garbage) in food waste create an additional processing concern for commercially available digesters, as non-organic contaminants can clog the digester and cause mechanical failures in moving parts within the sealed container.

The final challenge that must be overcome is the water balance within the digester. Food waste is approximately thirty percent solids. Traditional digestion technology requires material of less than ten percent solids fraction for both physical flow and biological processing. Addressing this challenge in commercially available technology requires adding large volumes of fresh water and consequently increased storage, treatment and disposal of additional liquid wastewater.

The Washington State Department of Ecology's Waste to Resources Program has funded research on anaerobic digestion at Washington State University for application to municipal organics recycling challenges, such as processing food waste. WSU's research has targeted systemic approaches and technology enhancements that attempt to resolve many of the barriers and deficiencies described above for commercially available technology.

The two most common commercially available approaches for digesting high solids fraction organic wastes like food waste are continuous two-stage wet (CTSW) and batch dry (BD). CTSW pre-treats the food waste to remove as many inert contaminants as possible; adds fresh and recycled water to dilute food wastes to less than 10 percent solids; and controls process pH and microbial inhibition by separating the digestion process into two steps (separate tanks). BD treats food waste at thirty percent solids without pre-treatment contaminant removal in piled batches in a vessel that continuously saturates the pile with a blend of fresh and recycled water to support growth of bacteria and control of the biological process<sup>17</sup>.

Each process has its own benefits and drawbacks. The dilution of CTSW creates a more favorable bacterial environment and biological treatment. However, the two phase system adds capital cost, expensive pre-treatment for necessary contaminant removal, and there is considerable amount of fresh water addition needed while generating a large amount of process liquid needing further treatment, storage and disposal. BD creates cost savings through avoided contaminant removal, theoretically uses less fresh water, and generates less process liquid. However, total biology activity is slowed reducing overall performance.

WSU's approach, which is applicable to either batch or continuous processes, separates the digestion process into two reactors. Food waste from 15 - 30 percent solids is digested in the first reactor (a "solids reactor") under controlled conditions to reduce inhibition while minimizing the impact of inert material. Liquids are removed in a recycle loop. The second reactor (a "liquid

reactor") rapidly digests and treats the liquid from the recycle loop and returns this treated liquid back to the solids digester. The overall effect of this two stage reactor system is to simultaneously enhance bacterial activity and process reaction rate, while controlling microbial inhibition and limiting fresh-water input. Outputs from the research in this report show that this system:

- (1) Stably digests food wastes with yields at least equal to CTSW systems;
- (2) Requires minimal plastics and garbage removal, comparable to BD systems;
- (3) Drastically reduces inputs of fresh water;
- (4) Reduces the need for storage and disposal of process liquid; and
- (5) Enables further recovery of ammonia and phosphorus concentrated bio-fertilizers.

While the two reactor approach and liquid by-product treatment systems add complexity and potential cost to the digestion system, it is anticipated that the increased process benefits of this modified approach will resolve many of the challenges of commercially available systems and will increase the likelihood of industry adoption of digestion for food waste treatment. WSU is currently working with industry partners to further refine and scale up the technology.

As has been proven in other industries, incremental technology and process improvements discovered through research and development can significantly improve commercial performance and contribute to overcoming technical, economic, environmental and social challenges to sustainability. While our research findings on municipal digestion do not resolve all the sustainability challenges facing municipal organics recycling, they are an important step in the right direction. Integrating anaerobic digestion with improved compost facility process controls *will enhance* organic recycling processing quality and performance, reduce odor, improve air quality, and reduce GHG emissions.

Further research underway could lead to breakthroughs in anaerobic digestion technology and may contribute to reducing process cost and increasing performance. Development of auxiliary technology, such as biogas conditioning and compression technology for generation of compressed methane fuel for use in waste management truck fleets, may create additional economic incentives for organics recycling facilities in our region to pursue anaerobic digestion as a key strategy for processing food waste.

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# Executive Summary

Waste 2 Resources (W2R) has strategically invested in research that will aid in the commercialization of “second-generation” organics processing technologies such as high solids anaerobic digestion that can better manage food waste and other materials that can be problematic for composting while enabling the recovery of energy and improving the capture and re-cycling of carbon and nutrients. In this case, composting and anaerobic digestion are complimentary in that the residual solids from anaerobic digestion are suitable for further processing via composting, with the combined technologies showing potential to reduce emissions, odors and greenhouse gas production.

The anaerobic digestion projects funded as part of this strategy represent two distinct approaches to promoting the commercialization of anaerobic digestion technologies necessary to catalyze transformative change in the organics management industry in Washington:

- A “problem-solving” strategy that focuses on making incremental changes to existing technologies that have not yet been adopted, in order to address adoption barriers that have been identified by the industry.
- A “pilot-project” approach that focuses on the development of new technologies.

The Center for Sustaining Agriculture and Natural Resources (CSANR), who managed this work, has successfully employed both approaches in prior organics processing work. Commercialization of research is challenging due to technical / scientific obstacles, financial constraints, policy barriers, and a myriad of additional socio-economic factors that affect the decision-making of enterprises considering technology adoption (i.e. risk tolerance, eligibility for public grants / loans, etc). Many excellent, technically and economically-proven technologies are never commercialized due to insurmountable barriers that are difficult to foresee. Thus, although the “pilot-project” approach to commercialization has been successfully implemented in many cases (and is more common than the “problem-solving” approach in the engineering field, particularly for “in-house” technology development conducted by large companies), it has a high risk of failure due to the many variables that cannot be controlled by inventors or investors. This is particularly the case in the context of limited research and development budgets, and for relatively expensive technologies such as anaerobic digestion.

To address these risks, W2R has invested in both the more risky, but transformative “pilot-project” approach, and a “problem-solving” or “technical support” approach. Under the “problem-solving” approach, the hypothesis is that enterprises with limited financial and technical resources are still fundamentally interested in new technology, but simply don’t have the risk tolerance, financial capacity, and / or financial flexibility to invest in or develop new technology. In addition, lending agencies are often unwilling to invest in unproven technologies. Because the adoption of new technology will improve the public well-being (i.e. by reducing environmental impact, promoting economic development, etc.) and because private sector financial institutions are generally fiscally conservative, it follows that investment of limited public resources in university research may provide the “tipping point” necessary to promote commercialization. In this approach, the needs assessment and collaboration process with

commercial enterprises is necessarily much more closely aligned, and the projected outcomes are more incremental than transformative, but the results of the research are much more readily adopted.

The two methods are targeted at different feedstocks (mixed food and green waste versus food waste alone), use very different processes (batch versus continuous), and are aimed at different operating conditions. Thus, the results from the two floor scale systems are not directly comparable, and the two approaches should be viewed as complimentary rather than competing.

## **Problem-solving approach: multi-reactor, liquid recycle system for high solids**

The overriding goal of this research project was to validate proof of concept of an integrated set of floor-scale leaching bed reactors (LBR), operated with a subsequent upflow anaerobic sludge blanket (UASB) reactor. In this design, each LBR is sequentially loaded with food/green material, and mixed with residual digested solids and leachate from a LBR that has completed its digestion cycle. Immediately after loading and sealing, a high flow rate of pre-digested, ammonia-stripped, high alkalinity UASB effluent is leached through the solids bed, accelerating hydrolysis and acidification and controlling pH near or above six, avoiding development of an acidified (phased) reactor. Once optimal leaching has been achieved and acidification is avoided, the LBR switches to an internal leachate recycle loop, using its own leachate to spur mass transfer, degradation kinetics, and accumulate alkalinity for additional stability. During this process, methanogens, either active since the original inoculant seeding or seeded from the bacteria present in the recycled, digested UASB effluent rapidly convert VFA to methane, while maintaining optimal low levels of acidic inhibitors and preparing the leachate solution for use within the next LBR. Products from LBR reactors are subsequently treated in a UASB reactor.

In line with the more incremental approach, this system builds upon previous work (Hofenk et al., 1984; ten Brummeler, 2000; Hedge and Pullammanappallil, 2007), incorporating subtle process changes and integrations that are based on review of literature and available science (Veeken and Hamelers, 2000; Hedge and Pullammanappallil, 2007; Speece, 1996; Cysneiros et al., 2008; Eastmann and Ferguson, 1981; Fox and Pohland, 1994; Frear et al., 2011). Key improvements included:

- Short duration, high flow rate leaching during initial stage.
- Non-sequential LBR and use of volumetrically low levels of seed.
- Non-phased, two-reactor approach.
- Phased insertion of bacterial populations.
- Nutrient recovery for removal of inhibitors, water savings and additional product revenue.

A floor-scale experimental system comprised of three parallel anaerobic reactor units was constructed and evaluated. Trials include both batch trials for internal LBR recycle alone and various forms of external feed from the UASB reactor, and trials of the full proposed system. While use of a single phase, dual reactor approach adds complexity and therefore has the potential to elevate capital and operating costs, floor-scale experimentation with the full design

(incorporating initial external recycle from the UASB to the LBR followed by internal recycling in the LBR) did show promise on several fronts compared to the simple batch, internal recycle-only baseline.

The batch LBR could be operated at sustained high pH levels, with minimal early depression, under a protocol of external leachate feed to the LBR during the initial days. These higher pH ranges promoted effective hydrolysis and acidification, and substantial methane production in both the LBR and the associated UASB reactor. Experimental evidence showed that the early stage external feed also positively impacted mass transfer and biodegradation kinetics.

Although resource limitations prevented full system optimization, data showed that external feed protocols with high alkalinity and moderate flow rate were preferred. In this experimental system, available alkalinities from the LBR/UASB system appeared to equilibrate at levels near 5 g/L CaCO<sub>3</sub>, making this the practical maximum to be fed to the system given the feedstock composition (a mixture of approximately 70% food waste with 30% mixture green grass and straw). Low feed rates, such as 1x the reactor liquid volume, did not allow for proper improvements in mass transfer from both a biological activity and product-inhibition removal perspective while too high of feed rates (i.e. 4x or above) presumably led to too much dilution, and washout of developing alkalinity and/or bacteria.

While other commercial versions of batch LBR digester systems recycle up to 50% or more of previously treated residual solids loaded as seed to new batches, this system operated effectively at 16.7% inclusion rates. Use of a primarily internal recycle regime allows for development of a final liquid batch high in bacterial seed and alkalinity which is ideal for new LBR loading and operation, contributing to a reduction in required residual solids for seed. Use of a greater seed ratio and/or changes to leaching operation (i.e. non-submerged, down-flow) could potentially improve upon degradation kinetics and allow for shorter solids retention times (SRT) within the LBR, though this hypothesis would need to be tested in future work.

Incorporation of the external feed also led to LBR products that were more preferred for subsequent treatment in the UASB reactor, in terms of the production of leachate volatile fatty acid (VFA) concentrations, VFA component profiles, and pH. The system's feed was at VFA concentrations near 3 g/L, comprised of a greater percentage of acetate and all at a pH near 7 (compared to leachate VFA concentrations bordering on 13 g/L, comprised of large percentages of butyric acid and at depressed pH (~4-5.5) for a similar system lacking external feed). Meanwhile, the use of external feed followed by internal recycle minimized the liquid recycle, pumping and UASB volume needs compared to continual external recycle. Although not optimized in this study, the length of external feed could be potentially reduced without fear of digestion upset, thereby further reducing the operating costs to the system.

The external feed LBR ceased hydrogen gas production sooner and had earlier production of methane compared to batch controls with no external feed. The small amounts of hydrogen gas generated could relatively easily be incorporated into the UASB stream and converted to methane, simplifying the number of end-products produced. By generating methane faster and to higher level (45.5% methane by day ten in the LBR reactor as compared 25.5% for internal recycle alone, with achievement of 68.7% by day twenty), the external feed system created a

single phase reactor system, with a feedback loop that further controlled pH and VFA concentrations and benefitted system stability.

Overall, performance data from the non-optimized system placed it right in the middle of expected performance as indicated by Hartmann and Ahring (2006) and well within desired targets. The full system had an organic loading rate (OLR) of 7.8 kg VS/m<sup>3</sup> day, yielding a specific methane productivity of 390 L CH<sub>4</sub>/kg VS loaded while reducing total volatile solids (VS) by 67% at a volumetric productivity of 1.8 m<sup>3</sup>/m<sup>3</sup> day.<sup>1</sup> Future optimization could improve system functioning. Use of a submerged up-flow system made necessary for testing at this scale, could have perhaps reduced mass transfer efficiency and resulted in abnormally long reaction times which in turn affected overall reaction time required and the longer-than expected timeframe for cessation of biogas production. These longer reaction times depress specific methane and volumetric biogas productivities; thus changes to this flow regime could conceivably raise the performance standards to the higher end of the expected range.

Results also indicate that much of the nitrogen has been converted to ammonia and is a concentrated in the reactor leachable liquids. Assuming 70% ammonia recovery efficiency, this translates to production of 400 MT of ammonia for facility that processed 100,000 MT/year. Importantly, integration of ammonia recovery would remove the concerns that ammonia could build to inhibitory levels over time, normally a concern in repeat batch operations such as this LBR system. Together with reductions in volatile solids, these data also indicate that AD pretreatment of food wastes in compost yards with this system could allow for further processing of solids residuals in compost piles with substantially reduced fear of ammonia or other volatile odorous compounds compared to composting alone. These conclusions point toward the full system providing important social, environmental and financial process improvements compared to a baseline internal-only leaching system.

Results from this floor-scale research suggest that the process could provide small but potentially significant improvement to the industry. Future research to follow up on this work should include the following key items:

- Optimizing scaled digestion technologies to the neutral pH, two reactor methane production concept, be it in batch, leaching-bed form as researched here or in other continuous fed systems as researched under the pilot project approach (described below).
- Integrating a nutrient recovery cycle within the two reactor system and verifying its concept and performance.
- Developing models for an integrated system to determine operating parameter ranges and impact on biomass growth and transfer.
- Funding, constructing and operating a 10% scale version of the identified process for next stage commercial development.

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<sup>1</sup> Compared to review literature that indicated that a high solids digester could operate at an OLR of near 10 kg VS/m<sup>3</sup> day and yield a specific methane productivity of 340 L CH<sub>4</sub>/kg VS loaded while reducing total VS by 60% at a volumetric productivity of between 1-4 m<sup>3</sup>/m<sup>3</sup> day)

## Pilot project approach: moderate solids digester system

Under the “pilot-project” approach, a floor-scale anaerobic system aimed at moderate solids (~15%) wastes such as food wastes was constructed and investigated. The moderate solids system included a patented high-solids digester previously developed at the laboratory scale, paired with a UASB reactor. Relatively high levels of flows are recycled from the second unit to the first unit, allowing effluent from the second reactor to seed the first reactor with methanogens, while retaining high levels of methanogens in the second stage digester. At the same time the pH can be maintained in a range that allows for methane production in both reactors.

Operations of the floor scale system indicated stable reactor conditions over a six month operational period with food waste feedstock, and incorporated two additional design improvements. First, an intermittent mixing strategy has proven sufficient to break up biowaste clumps (thick, aggregated clumps of food wastes) if the system utilizes two additional passive mixing forces: a design that sprays the effluent from UASB seed reactor into the high solids digester at the top inlet, and natural mixing arising from the force of biogas bubbles rising to the top of the high solids digester. While these strategies have been applied in wastewater treatment plants, they have not previously been successfully applied for a moderate-solids system such as the one proposed here. Second, a liquid-solid separation technology that allows solids to rise to the top and removes liquid from lower layers takes advantage of natural forces of the ongoing anaerobic digestion was utilized. Together these improve the potential to cut mixing costs by nearly 60% by simplifying the system, as calculated by previous laboratory-scale research (Zaher et al., 2009).

The high solids digester was operated at a pH of 6.0-7.0 (optimal for volatile fatty acid production and methanogen growth). The system had an organic loading rate (OLR) of 5.0 kg VS/m<sup>3</sup> day, yielding a specific methane productivity of 290 L CH<sub>4</sub>/kg VS loaded while reducing total volatile solids (VS) by 48.63% at a volumetric productivity of 2.0 m<sup>3</sup>/m<sup>3</sup> day. This experimental system was not fully optimized, and thus further performance gains may be achievable.

In addition to improvements, operation, and testing of the system at floor-scale, a process model, based on anaerobic digestion model No. 1 (ADM1) was developed and validated based on experimental data. Once validated, the model was used to examine the impact of changing the recirculation rate on process stability and kinetics, since this information is difficult to obtain from experiments. Model results suggested that a recirculation rate of 0.19 m<sup>3</sup>/day was optimal, maintaining pH at the range of 7.0-8.0 in the high solids digester. Under these conditions, the pH in the UASB seed reactor was maintained at 7.5-8.0 and most of VFA was consumed in this reactor.

Lastly, effluent from the floor-scale system was successfully treated in a laboratory scale integrated, multifunction ammonia removal and nutrient recovery system (IMARNRS) previously developed at WSU (Jiang et al., 2009). Results showed 93% total ammonia nitrogen

removal achieved in one day, and phosphate removal of 45% in six hours. Successful integration with nutrient removal will improve system stability by removing N and P that could inhibit the system. Using experimental data, it was estimated that integrating a nutrient removal and recovery system into the effluent recycle loop of a facility processing 100,000 MT of food waste per year could produce 376 MT of nitrogen and 40 MT of phosphorus per year.

In addition to the successful design improvements incorporated at floor scale, experience with the floor scale project generated several lessons that should be applied future work at the pilot scale. These include introducing biogas from the high solids digester into the UASB during initial system set-up, the use of a vertical liquid outlet in the high solids digester, and simplification of the system through the use gravity transport (rather than pumps) where feasible.

The results presented here support ongoing work towards scale-up of the new HSAD system. To achieve successful commercialization of this technology, collaboration with an industry partner is needed to implement the system at the pilot scale. Goals of the pilot scale design would be to refine the engineering design and generate additional economic analysis relevant for commercial scale planning.

# Part I: Multi-Reactor Liquid Recycle System for High Solids

This project, one of two anaerobic digestion research projects carried out by Washington State University (WSU) for the Washington Department of Ecology (Ecology) Waste 2 Resources (W2R) program, constructed and investigated a floor-scale anaerobic digestion system for high solids (~30%) combined food waste and green waste. Ecology estimated that in 2008/2009, Washington State, with a diversion rate of 55%, landfilled nearly 5 million tons of MSW, with about 18% by weight being food scraps, and 4% leaves and grass (Washington Ecology 2010a, 201b).

The overarching goal of this project was to validate proof of concept of an integrated set of floor-scale leaching bed reactors (LBR), operated with a subsequent upflow anaerobic sludge blanket (UASB) reactor. Such a design would be highly appropriate for integration with existing composting facilities, and aimed to alleviate concerns with odor release, compost quality, and processing times that have come up as composting facilities have accepted more highly degradable food wastes. Specific objectives included:

- Test the batch, internal recycle of LBR for phased digestion in a single reactor to determine effectiveness and whether or not the process requires an external loop leaching to operate effectively.
- Test the impacts of high-rate, short-duration external leaching of LBR on controlling initial pH and acid production. Investigate factors of leachate rate and alkalinity concentration against pH, VFA, SCOD, biogas, methane content, and hydrogen content indicators.
- Determine the theoretical methane potential, energy production and nutrient recovery from test feedstock, a mixture of cafeteria food scraps and green waste;
- Assess yields and productivities of the full, proposed system.

In this design, each LBR is sequentially loaded with food/green material, and mixed with residual digested solids and leachate from a LBR that has completed its digestion cycle. Immediately after loading and sealing, a high flow rate of pre-digested, ammonia-stripped, high alkalinity UASB effluent is leached through the solids bed, accelerating hydrolysis and acidification and controlling pH near or above six, avoiding development of an acidified (phased) reactor. Once optimal leaching has been achieved and acidification is avoided, the LBR switches to an internal leachate recycle loop, using its own leachate to spur mass transfer, degradation kinetics, and accumulate alkalinity for additional stability. During this process, methanogens, either active since the original inoculant seeding or seeded from the bacteria present in the recycled, digested UASB effluent rapidly convert VFA to methane, while maintaining optimal low levels of acidic inhibitors and preparing the leachate solution for use within the next LBR. Products from LBR reactors are subsequently treated in a UASB reactor.

This system combined and built upon previous work carried out by others in the field (Hofenk et

al., 1984; ten Brummeler, 2000; Hedge and Pullammanappallil, 2007). Process changes and integrations used in the design were based on review of literature and available science (Veeken and Hamelers, 2000; Hedge and Pullammanappallil, 2007; Speece, 1996; Cysneiros et al., 2008; Eastmann and Ferguson, 1981; Fox and Pohland, 1994; Frear et al., 2011). Key features included:

- Short duration, high flow rate leaching during initial stage.
- Non-sequential LBR and use of volumetrically low levels of seed.
- Non-phased, two-reactor approach.
- Phased insertion of bacterial populations.
- Nutrient recovery for removal of inhibitors, water savings and additional product revenue.

A floor-scale experimental system comprised of three parallel anaerobic reactor units was constructed and evaluated. Trials include both batch trials for internal LBR recycle alone and various forms of external feed from the UASB reactor, and trials of the full proposed system. While use of a single phase, dual reactor approach adds complexity and therefore has the potential to elevate capital and operating costs, floor-scale experimentation with the full design (incorporating initial external recycle from the UASB to the LBR followed by internal recycling in the LBR) did show promise on several fronts compared to the simple batch, internal recycle-only baseline.

The batch LBR could be operated at sustained high pH levels, with minimal early depression, under a protocol of external leachate feed to the LBR during the initial days. These higher pH ranges promoted effective hydrolysis and acidification, and substantial methane production in both the LBR and the associated UASB reactor. Experimental evidence showed that the early stage external feed also positively impacted mass transfer and biodegradation kinetics.

Although resource limitations prevented full system optimization, data showed that external feed protocols with high alkalinity and moderate flow rate were preferred. In this experimental system, available碱性ities from the LBR/UASB system appeared to equilibrate at levels near 5 g/L CaCO<sub>3</sub>, making this the practical maximum to be fed to the system given the feedstock composition (a mixture of approximately 70% food waste with 30% mixture green grass and straw). Low feed rates, such as 1x the reactor liquid volume, did not allow for proper improvements in mass transfer from both a biological activity and product-inhibition removal perspective while too high of feed rates (i.e. 4x or above) presumably led to too much dilution, and washout of developing alkalinity and/or bacteria.

While other commercial versions of batch LBR digester systems recycle up to 50% or more of previously treated residual solids loaded as seed to new batches, this system operated effectively at 16.7% inclusion rates. Use of a primarily internal recycle regime allows for development of a final liquid batch high in bacterial seed and alkalinity which is ideal for new LBR loading and operation, contributing to a reduction in required residual solids for seed. Use of a greater seed ratio and/or changes to leaching operation (i.e. non-submerged, down-flow) could potentially improve upon degradation kinetics and allow for shorter solids retention times (SRT) within the LBR, though this hypothesis would need to be tested in future work.

Incorporation of the external feed also led to LBR products that were more preferred for subsequent treatment in the UASB reactor, in terms of the production of leachate volatile fatty acid

(VFA) concentrations, VFA component profiles, and pH. The system's feed was at VFA concentrations near 3 g/L, comprised of a greater percentage of acetate and all at a pH near 7 (compared to leachate VFA concentrations bordering on 13 g/L, comprised of large percentages of butyric acid and at depressed pH (~4-5.5) for a similar system lacking external feed). Meanwhile, the use of external feed followed by internal recycle minimized the liquid recycle, pumping and UASB volume needs compared to continual external recycle. Although not optimized in this study, the length of external feed could be potentially reduced without fear of digestion upset, thereby further reducing the operating costs to the system.

The external feed LBR ceased hydrogen gas production sooner and had earlier production of methane compared to batch controls with no external feed. The small amounts of hydrogen gas generated could relatively easily be incorporated into the UASB stream and converted to methane, simplifying the number of end-products produced. By generating methane faster and to higher level (45.5% methane by day ten in the LBR reactor as compared 25.5% for internal recycle alone, with achievement of 68.7% by day twenty), the external feed system created a single phase reactor system, with a feedback loop that further controlled pH and VFA concentrations and benefitted system stability.

Overall, performance data placed the system right in the middle of expected performance as indicated by Hartmann and Ahring (2006) and well within desired targets. The full system had an organic loading rate (OLR) of 7.8 kg VS/m<sup>3</sup> day, yielding a specific methane productivity of 390 L CH<sub>4</sub>/kg VS loaded while reducing total volatile solids (VS) by 67% at a volumetric productivity of 1.8 m<sup>3</sup>/m<sup>3</sup> day.<sup>2</sup> Use of a submerged up-flow system made necessary for testing at this scale, could have perhaps reduced mass transfer efficiency and resulted in abnormally long reaction times which in turn affected overall reaction time required and the longer-than expected timeframe for cessation of biogas production. These longer reaction times depress specific methane and volumetric biogas productivities; thus changes to this flow regime could conceivably raise the performance standards to the higher end of the expected range.

Results also indicate that much of the nitrogen has been converted to ammonia and is concentrated in the reactor leachable liquids. Assuming 70% ammonia recovery efficiency, this translates to production of 400 MT of ammonia for a facility that processes 100,000 MT/year. Importantly, integration of ammonia recovery would remove the concerns that ammonia could build to inhibitory levels over time, normally a concern in repeat batch operations such as this LBR system. Together with reductions in volatile solids, these data also indicate that AD pretreatment of food wastes in compost yards with this system could allow for further processing of solids residuals in compost piles with substantially reduced fear of ammonia or other volatile odorous compounds compared to composting alone. These conclusions point toward the full system providing important social, environmental and financial process improvements compared to a baseline internal-only leaching system.

Results from this floor-scale research suggest that the process could provide small but potentially significant improvement to the industry. Future research needs needed to follow up on this work include the following key items:

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<sup>2</sup> Compared to review literature that indicated that a high solids digester could operate at an OLR of near 10 kg VS/m<sup>3</sup> day and yield a specific methane productivity of 340 L CH<sub>4</sub>/kg VS loaded while reducing total VS by 60% at a volumetric productivity of between 1-4 m<sup>3</sup>/m<sup>3</sup> day)

- Optimizing scaled digestion technologies to the neutral pH, two reactor methane production concept, be it in batch, leaching-bed form as researched here or in other continuous fed systems as researched under the pilot project approach (described below).
- Integrating a nutrient recovery cycle within the two reactor system and verifying its concept and performance.
- Developing models for an integrated system to determine operating parameter ranges and impact on biomass growth and transfer.
- Funding, constructing and operating a 10% scale version of the identified process for next stage commercial development.

# **Section 1: Concerns with composting food scraps and fresh grass**

## **Background**

Communities in the US are encouraging per capita reductions in municipal solid waste (MSW) generation due to concerns about landfill operations and sustainability (WA Ecology 2010b). This includes implementing strategies to increase rates of waste recycling and diversion (WA Ecology, 2010b). A US EPA (2008) summary of waste generation shows that in 2007 54% of the 230 million metric tons (MMT) of MSW produced in the US was disposed of in landfills.

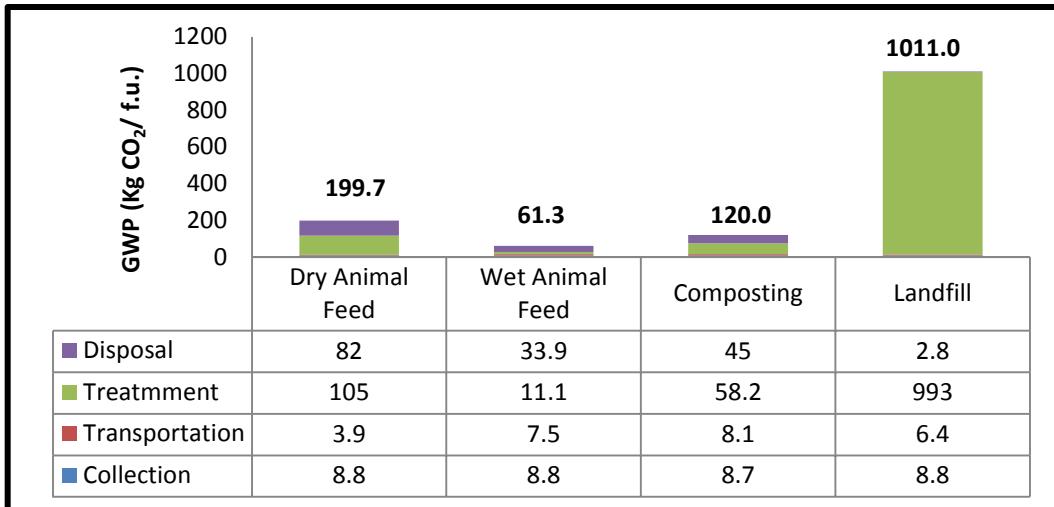
Comparatively, in 2009, Washington State landfilled about 45%, or nearly 5 million tons of MSW (WA Ecology 2010b).

While active recycling and diversion projects exist for many categories of MSW, food scraps and yard trimmings are receiving particular interest. Representing 21.1% of total MSW production nationally, yard trimmings have had a long history of being recycled and treated through aerobic composting. The US EPA (2008) estimated that over 64.1% of total production was diverted for treatment and eventual production of soil amendment products. In contrast, they estimated that only 2.6% of the 28.8 MMT annually produced (12.5% of total MSW) were recycled and treated, leaving the overwhelming majority decomposing in landfills (US EPA, 2008) (see Figure 1). In Washington State, landfilled wastes contained nearly 55% organics, about 18% food waste, and 4% leaves and grass (WA Ecology, 2010a; WA Ecology 2010b).



Figure 1. Roosevelt Regional landfill Washington State (photo: WA Ecology, 2010b).

This current disposal strategy can have negative impacts on the climate and other environmental indicators. Using a life cycle analysis (LCA), Kim and Kim (2010) demonstrated that the cradle to grave global warming potential (GWP) of landfill disposal of food scraps, even assuming partial recovery and utilization of methane, is nearly eight times higher than composting or animal feed diversion, two competing treatment options (see Figure 2). Using a similar LCA methodology but looking at a broader range of indicators, Lundie and Peters (2005) showed that in-sink food waste processing and centralized composting out-performed landfill disposal of food wastes in terms of key environmental indicators of human, aquatic, and terrestrial toxicity, and acidification and eutrophication potential.



GWP = global warming potential (measured as kg CO<sub>2</sub>/functional unit, where functional unit is defined as the by-products from dry feeding, wet feeding, composting, and landfilling of 1 MT of food wastes).

Figure 2. LCA for various food scrap treatment options (Kim and Kim, 2010).

Numerous countries are recognizing the societal and environmental benefits of landfill diversion and are setting regulations encouraging or requiring the recovery of organics – with a recent focus on food scraps (Levis et al., 2010). South Korea outlawed direct landfill disposal of raw food scraps in 2005 and achieved recycling rates of over 94% recycling in 2006. Of the recycled food waste, 45.2%, 44.9% and 9.8% of the material was used for animal feed, composting, and renewable energy via anaerobic digestion (AD), respectively. Despite the impressive recovery rates, the sudden, legislation was not without its concerns. Limitations in commercially-available conversion processes resulted in wastewaters from treatment options often finding their way to open-ocean dumping while residual solids were ultimately incinerated or land-filled (Kim and Kim, 2010).

While far behind other nations, Washington State is ahead of other US states in regard to the recycling of food scraps. The state composts nearly 8% of the 977,000 tons annually produced at 17 separate sites (WA Ecology, 2010b), and ongoing community interest is likely to spur continued growth in food scrap recovery rates. However, with increased recovery, Washington State and other areas have experienced a corresponding increase in processing burden on existing infrastructure, and concomitant concerns about odor generation from compost facilities and quality of compost product (WA Ecology, 2010b; Kim and Kim 2010; Sundberg et al., 2011).

## Compost of food scraps and fresh green waste

Compost is an actively commercialized treatment process for the organic fraction of municipal solid waste (OFMSW). It can result in mass reductions of 55-65% dry weight for grasses, mixed yard waste and food scraps (EPA, 2003). This yields a stabilized, marketable soil amendment that improves soils, by increasing water and air holding capacity, and by slow release of macro- and micro-nutrients (Vaughan and Malcolm, 1987). Optimization of the degradation process is controlled by numerous factors including waste type, carbon availability, C/N ratio, temperature, aeration rate, moisture content, pH, and percent and type of bulking materials utilized (Eiland et al., 2001; Barrington et al., 2002; Adhikari et al., 2009). Prompted by increased interest in diversion of food scraps from landfills, as well as ongoing challenges with extreme seasonal

fluctuations in organic waste types (e.g. high spring volumes of grass clippings), particular attention has recently been placed on understanding the intricacies of composting material composed primarily of fresh grasses and/or food scraps (Beck-Friss et al., 1999; Chang et al., 2006; Richart and Walker, 2006; Eklind et al., 2007; Cheung et al., 2010; Sundberg et al., 2011). While grasses and food scraps can vary widely in their composition based on their source, collection, and/or season, some common features can be summarized (Table 1).

Table 1. Characteristics of food scraps and fresh green waste.

Parameter	Food Scraps Range	Fresh Green Waste Range
Moisture (%)	66-81	15-41
pH	4.1-6.1	5.9-7.4
Fixed Solids/Ash (% DM)	6-15	3-6
Total Carbon (% DM)	33-54	33-57
Total Nitrogen (% DM)	1.1-3.1	0.6-1.6
C/N Ratio	14-30	30-56
Volatile Fatty Acids (VFA) (% DM)	0.5-5	0.25-0.34

Source: Sundberg et al., 2011; Kim and Kim, 2010; Adhikari et al., 2009; Eklind et al., 2007; Zhang et al., 2007; Lopez et al., 2010, Mahnert, 2005

Compost operators are trained to maintain moisture and C/N ratios within preferred ranges (Diaz et al., 1993). However, in composting food waste, additional focus should be placed on the low pH values. The acid range pH in Table 1 shows that materials arrive at the facility gate partly decomposed, containing volatile fatty acids (VFA) and other odor causing chemicals, including ammonia, sulfur and volatile organic compounds (VOC). Each of these can cause immediate odor and air/water quality impacts during preprocessing or compost processing, curing and storage. Figure 3 outlines the step-by-step flow process on a typical compost facility, pointing to specific places where material handling and/or biological processing could lead to odor and air/water quality concerns.

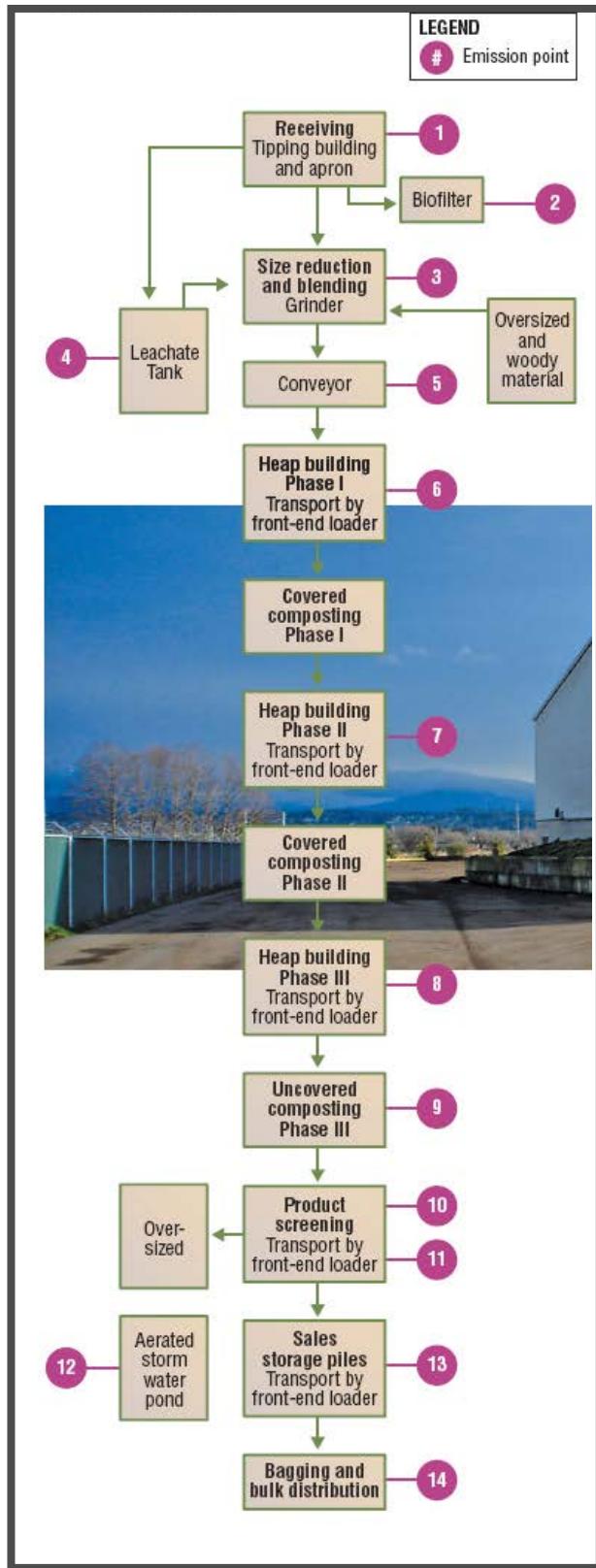


Figure 3. Compost facility schematic (Biocycle, 2008).

Many compost facilities have adopted protocols to control odor being released from putrescent materials, including walled and covered processing buildings and specific materials handling procedures. However, even when these control measures are implemented, pH and VFA can have a negative biological effect on the aerobic compost process itself. Initial low pH has been shown to severely inhibit aerobic organisms within compost piles (Sundberg and Jonsson, 2008; Kurola et al., 2010; Partanen et al., 2010). Acidic conditions result in low concentrations of *Bacillales* and Actinobacteria which are critical for effective composting, particularly during initial staging (Cheung et al., 2010). Low pH also promotes communities of *Lactobacillus* and *Escherichia coli* bacteria (Partanen et al., 2010, Smars et al., 2002; Sundberg et al., 2004), which can rise to high concentrations—above those of the preferred organisms—due to their absolute and comparative tolerance of heat and acidic conditions. At low pH (<6) and high temperature (>40 °C) *Lactobacillus* and *Escherichia coli* bacteria can proliferate and produce VFA (products of their metabolism). This causes further depression of the pH and can result in a feedback loop, leading to a sustained and problematic inhibition of beneficial microbes (Brinton, 1998; Smars et al., 2002; Sundberg and Jonsson, 2005; Cheung et al., 2010; Sundberg et al., 2011). Anaerobic microbes within the compost piles may exacerbate the problem. This has been found at large compost operations with high flow rates and excessively large piles (Brinton 1998; Reinhardt 2002). Under depleted oxygen conditions, the anaerobic cultures are also capable of producing VFA and lowering the pH, further enriching the negative feedback cycle.

When this feedback loop of high heat and acid occurs during the early stages of composting, it can result in abnormally long first stages as well as a lengthened timeline for final compost maturation. The extended kinetics does not necessarily yield greater mass releases of volatiles, but it does extend the release over a greater period of time. It may also extend regularly scheduled movements and turnings of piles, leading to a potential for greater release of volatiles. For example, typical compost feeds, including food scraps and fresh grasses, produce sequential CO<sub>2</sub> and ammonia peaks during the 3-8 and 12-40 day periods, respectively. VOCs are also released primarily during this early period, subsequently tapering to lower values with an approximate first order rate constant (with the exception of ketones) (EPA, 2003). Thus, an operator may assume that at the end of the 30-40-day cycle, piles could be uncovered, transferred, and turned with minimal release of volatile ammonia and VOC. However, if delays occur due to acidification and elevated temperatures, this assumption could be false. This is particularly true in regard to food waste compost (Figure 4), which releases nearly 65% of its initial N as ammonia-N (food waste emits 34-41 g N/dry kg which is about 20x that of other common compost feeds; EPA, 2003).

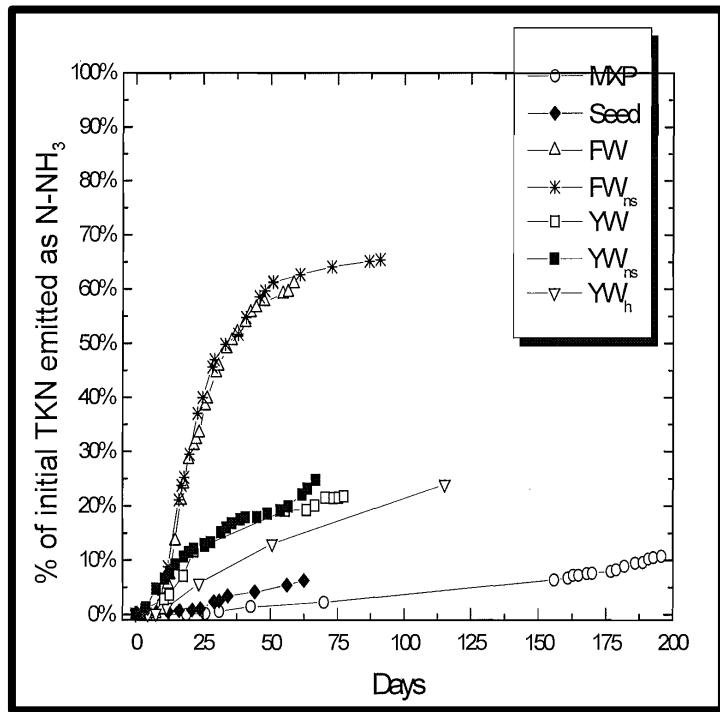


Figure 4. Evolution of ammonia-N as percent of initial total N (MXP: mixed; Seed: aged compost used as initial seed; FW: food waste with no-seed; FW<sub>ns</sub>: food waste without seed; YW: yard waste with seed; YW<sub>ns</sub>: yard waste without seed; YWh: household yard waste; EPA, 2003).

While ammonia is released in much larger volumes than VOC, VOC do represent their own particular odor and health concerns. Primary VOC resulting from grasses and food scraps are terpenes, alcohols, acids, esters of acids, organic-sulfur compounds, and ketones. Significant levels of dimethyl disulfides, dimethyl trisulfides and limonene were detected as well for food scraps (EPA, 2003), with total VOC emissions for a specific class of compounds being 2.8 and 2.1 mg VOC/dry weight for food scraps and grasses, respectively (EPA, 2003).

In addition to increased risk of odor releases, delays in the compost process affect facility economics and product stability. Extended composting time leads to reduction in available composting space within the facility which in turn either leads to less product flow-through (negatively impacting economics) or a need to compensate for longer times with larger piles (reducing product quality). Since large compost facilities are already impacted by depressed sale prices and markets for finished compost (EPA, 2003), the presumed choice for many facilities will be maintenance of existing flows through the use of larger piles and/or higher thermophilic operating temperatures.

Larger piles significantly increase the risk of producing anaerobic environments and result in pile conditions that only strengthen the negative feedback that already exists regarding acids, pH and longer retention times. In addition, anaerobic processing yields significantly greater levels of climate changing greenhouse gases (GHG) such as methane (CH<sub>4</sub>) and nitrous oxide (N<sub>2</sub>O) (Hellman, 1997). Beck-Friis et al. (1999) showed that mean CH<sub>4</sub> and N<sub>2</sub>O fluxes for large commercial piles were 35 g CH<sub>4</sub>/m<sup>2</sup> day and 0.261 g N<sub>2</sub>O/m<sup>2</sup> day, respectively, each 4x and 3x

that of small piles, respectively. Lastly, large piles can result in inferior product quality that at times does not meet federal regulations regarding pathogen controls. An extensive sampling of 94 non-sludge commercial compost products by Brinton et al. (2009) shows that 20%, 28% and 47% of products sampled exceeded regulator limits for *Clostridium perfringens*, *fecal Coliforms*, and *fecal Streptococci*, respectively.

Large piles for the most part result in a higher operating temperatures. Richert and Walker (2006) found that temperatures can be as high 80-90 °C within many facilities, particularly those utilizing active covers. This is significantly above the 52-60 °C temperatures considered to be optimal. Higher temperatures could be controlled with increased aeration rates, but the cost of aeration as well as difficulties in regulating such flows lead many facilities to opt for higher, non-ideal temperatures, particularly if the increased temperature is *assumed* to enhance kinetic degradation and allow the facility to achieve high feedstock throughput (Eklind et al., 2007). From the perspective of volatile emissions and decomposition rates, this choice can be detrimental. Eklind et al. (2007) showed that thermophilic temperatures greater than 67 °C lead to more than double the ammonia emissions as compared to lower temperatures of 40 °C and 55 °C, degrading the resulting compost fertilizer quality. Similarly, Eklind et al. (2007) and Diaz et al. (1993) found that optimal degradation occurred not at higher temperatures (67 °C or greater) as was assumed, but at the more moderate and previously described 55 °C level, with temperatures > 70 °C being shown to significantly inhibit degradation. Worse, the above studies were completed at constant aeration rates, but higher aeration rates are known to be required for commercial operations using large piles. Higher aeration rates would only make the ammonia and volatiles emission rate higher, as aeration rate is directly correlated to ammonia release (Elwell et al., 2002).

These examples indicate that balancing important environmental and economic indicators is important to identify a preferred composting temperature for food scraps and/or green waste. A temperature range near 52-55 °C can maximize degradation while minimizing ammonia emissions. Incorporating the earlier concepts related to pH, VFA and acid forming bacteria, it may be preferred to target an initial low temperature (<40 °C) prior to temperature elevation in later stages for highly biodegradable organic material such as food scraps (Table 2).

Table 2. Preferred operating conditions for compost of low pH, high VFA food scrap/fresh grasses containing bio-waste.

	Preferred Operation	Reasons
Pile Size	Medium	Reduce anaerobic zones, temperature control (Beck Friis et al., 1999)
Phased Temperature	37 °C initial until pH > 6	VFA, pH control, removal of inhibition feedback loop (Sundberg Jonsson, 2008)
Compost Temperature	52-55 °C	Degradation kinetics, pathogen destruction, ammonia loss (Richart and Walker, 2006, Eklind et al., 2007)
Aeration Rate	≤ 15% DO	High degradation, minimize ammonia loss, control temp and cost (de Bertoldi et al., 1988)
Moisture Content	52-60%	Control anaerobic, allow aerobic kinetics (Schulze, 1961)
Seed	Yes/No	Improves performance (Grey et al., 1971, but not very important with grass and food (EPA, 2003)

### Incorporation of anaerobic digestion with compost

Composting food scraps and/or fresh green grasses can be problematic in large-scale facilities though the discussion above illustrates how relevant science could be used to optimize degradation and enhance commercial economics. Particular problem areas identified include:

- Low pH, high VFA characteristics of already decomposing material entering compost facilities.
- Release of ammonia and VOC during processing especially during delays in processing, turning and curing induced by negative feedback loop.
- Production of higher levels of GHGs while using large piles.
- Negative effect on pathogen control.
- Negative effect on facility economics through impacts to product quality, energy inputs, mass flows, etc.
- Loss of final agricultural highest fertility product value by ammonia volatilization.

A cursory analysis of the functions of anaerobic digestion (AD) indicates that a combination of AD with compost could synergistically address these concerns. AD is a biological process that converts VFA into methane-rich biogas through the use of *archae* methanogens while also raising pH of resulting liquids and solids through development of high alkalinity and buffer systems. Ammonium is produced from organic nitrogen in an anaerobic digester. In this confined system the ammonia can be easily volatized and collected, decreasing N emissions and resulting in fertilizer sales. Methane emissions occurring during the digestion process are used to produce renewable energy (whereas they are normally lost as GHG emissions in compost). This can offset the energy requirements of the compost process, estimated at 25.2 kWh/ton processed for static aerated piles (de Bertoldi et al., 1983). Additional economic benefits can result from the 60% mass reductions occurring during AD as a large fraction of VOC is converted to methane (Hartmann and Ahring, 2006) resulting in more efficient use of compost yard space. Lastly, AD with its known temperature and time treatment allows for a primary as well as secondary (if secondary compost included) treatment process for control of pathogens.

Anaerobic digestion is not without its drawbacks. Inclusion of AD into an AD/compost combined process increases capital and operating expenses. Thus, additional techno-economic analysis of the combined AD-compost system is warranted. Early analyses of food scrap AD standalone systems show a 0.018 KW/wet ton net electrical production or \$14/wet ton electrical revenue capability (Banks et al., 2011). The conversion of VFA and volatile solids with highly biodegradable food scraps and fresh green waste can be prone to run-away kinetics during the first stages of AD. If not controlled, this can result in a lowering of pH and subsequent souring, failing or extended recovery/treatment time, not unlike the negative feedback loop described for composting. In AD applications that do not incorporate N recovery, ammonia can also be a concern, with a danger that it reaches concentrations which inhibit digestion.

Evidence also shows that the relationship between AD and compost for final treatment of AD residual solids can be complex. Mata-Alvarez et al. (2000) have shown that VOC and ammonia emissions from a compost operation utilizing primarily food waste can be lowered from a mean of 747 g/MT to a mean of 101 g/MT through the use of a sequential AD/compost operation. Drennan and Distefano (2010) showed that subsequent aerobic composting of residual AD solids achieved low levels of VOC and in particular VSC (volatile sulfur compounds) emissions after only 10-15 and 15-20 days, respectively, producing a solid product that met US Compost Council guidelines with respect to many indicating factors, including oxygen uptake rates. In this study compost operating temperatures, were typical of late stage compost systems, achieving only moderate mesophilic temperatures before stabilizing at lower ambient temperatures.

## Section 2: Anaerobic digestion of food scraps and green waste

### Background

This work investigates new anaerobic digestion methods suitable for incorporation with composting operations and identifies operating parameters for recovering energy and fertilizers from highly degradable food and green waste. Appropriately operated digesters can effectively treat high organic waste loads, but are not a panacea. Operated beyond parameters, microbial activity in digesters can be severely inhibited by the high biodegradability, low pH, and high nitrogen levels present within food and green waste.

These sensitivities to feedstock composition are due to the fact that AD is a process of sequential and synergetic biological steps that utilize a consortium of microorganisms with varying growth and substrate uptake rates. Various microorganisms carry out the four stages that make up digestion: hydrolysis (where particulates are solubilized and large polymers converted into simpler monomers), acidogenesis (where simple monomers are converted into volatile fatty acids), acetogenesis (where volatile fatty acids are converted into acetic acid, carbon dioxide, and hydrogen), and methanogenesis (where acetates are converted into methane and carbon dioxide, while hydrogen is consumed) (Figure 5). When this process is managed well, it can readily digest organic waste and subsequent intermediary metabolites without fear of inhibition. If not managed, the initial microbial decomposition steps of hydrolysis and acidification yield products that may overwhelm the final step of methane-formation, that may depress overall digester performance or sour and destroy the microbiological process.

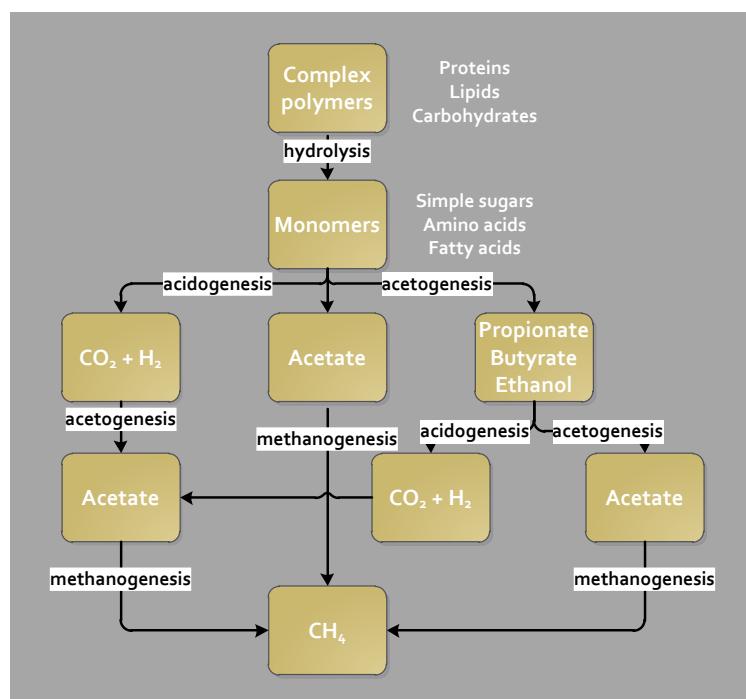


Figure 5. Anaerobic digestion process (Bitton, 1994).

This section provides some context for WSU's work on anaerobic digestion of food scraps and green waste, focusing on project economics in the US, the various technologies currently used for high solids organic wastes,

## **Anaerobic digestion economics**

A concern often mentioned is the high capital and operating costs that have traditionally been associated with AD units, as they have developed over the last decade in Europe (Figure 6).

Figure 6 indicates that a typical 100,000 MT plant would have capital costs around \$15-20 million with operating costs at \$80-100/MT treated (considering only electrical production, which many investors focus on). These costs might be economically viable for projects capable of receiving 20-30¢/kWh, which are typical in Western Europe. However, within the US, typical electrical sales prices are near 9¢/kWh. At this price, the project payback period would be near 20 years—too long for most project investors.

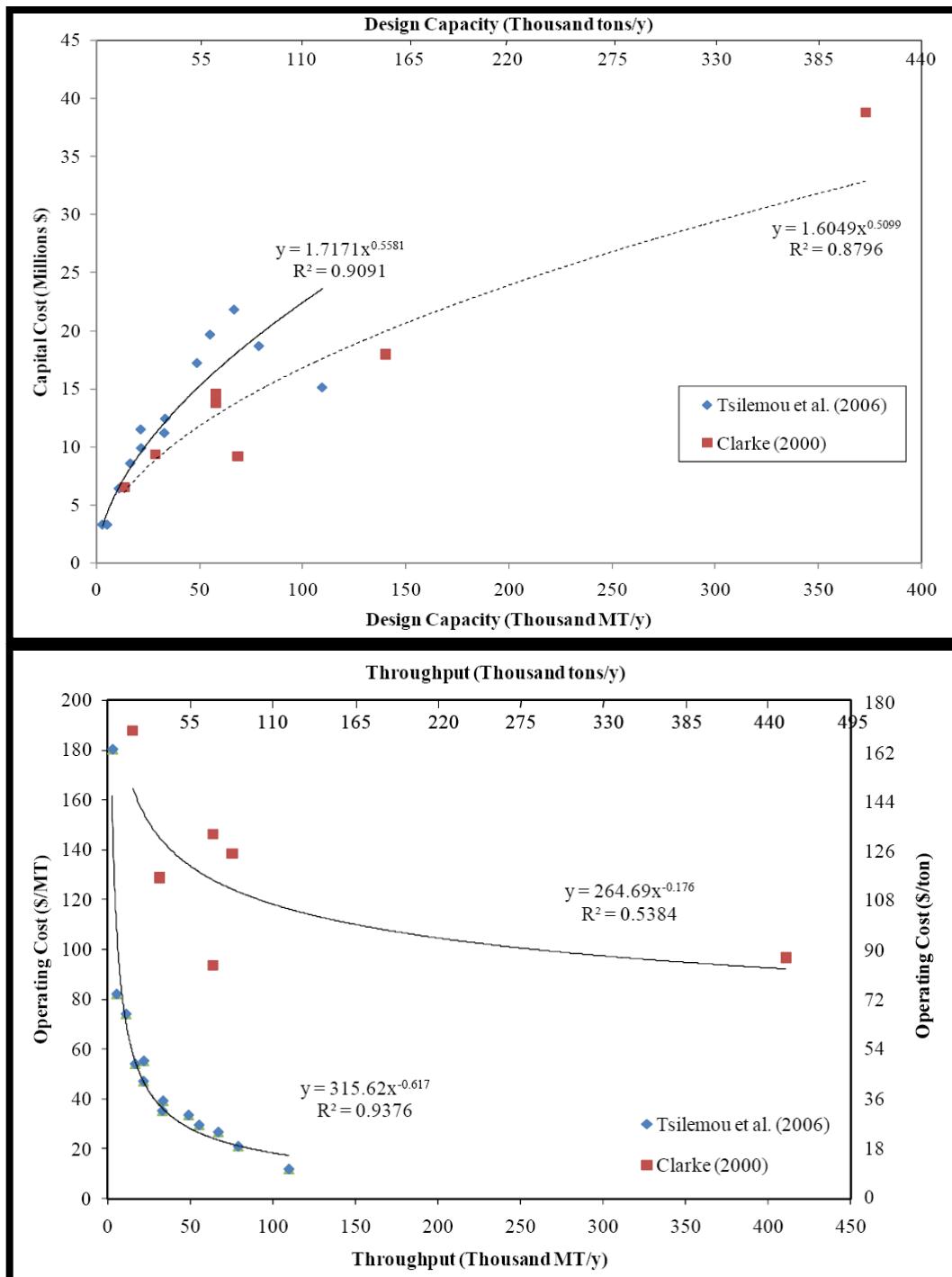


Figure 6. Capital cost by design capacity and operating costs by flow throughput for commercial OFMSW digesters as reported in Hartmann and Ahring (2006).

Clearly, in the US, project economics based solely on electricity production are not favorable. A number of strategies, including reductions in construction and operating costs, development of higher value and/or additional product revenue streams, providing extended treatment in tandem with a compost facility, increasing the mass and quality of compost, or a combination of these

strategies may create return on investments that are more favorable. To find a way forward, HSAD digester technology may be able to learn lessons from farm-based, manure-fed digesters. These digesters, despite revenue pricing constraints (i.e. low electrical sale prices), have developed a steady number of viable projects in the US (>150 farm-based AD projects at an installation rate of 15/year for the last ten years) from a combination of (1) technologies that are primarily less costly in operation, if not capital construction; (2) multiple product revenue streams being developed; and (3) technologies and product streams being able to resolve regulatory concerns while producing renewable energy<sup>3</sup> (Frear et al., 2011; Bishop and Shumway, 2009).

## **Technologies used for anaerobic digestion of food waste and green waste**

The literature is laden with different descriptive phrases that attempt to correctly explain the various digestion technologies used for solid waste organics AD, i.e. wet, dry, high solids, etc. Many terminologies use words such as “high solids” or “dry”, indicating that solid waste organics, particularly food scrap and green waste, tend to be higher in solids concentration (15-35% total solids (TS)) than is experienced in other common AD fields such as municipal wastewater and animal manure treatment (2-10% TS). Unfortunately, these words can contribute to misunderstandings because during the digestion process, the initial feedstocks may be diluted in various ways and thus might not be particularly “dry” or “high solids”.

For simplicity and future reference in this report, Figure 7 outlines the major types of technologies, using new terminology aimed at removing potential misunderstandings and more directly describing the processing technologies. Each of the approaches is capable of being operated at either mesophilic (20 to 45 °C) or thermophilic (45 to 70 °C) temperature regimes. The choice of temperature regime will not be the focus of this summary section or research project, as this area is still disputed. While some literature has suggested that thermophilic temperatures enhance overall kinetics, volumetric productivity, and even specific methane productivity while operating stably (Hartmann and Ahring, 2006), other researchers still have lingering questions regarding some aspects of operational stability (Mata-Alvaraz, 2000).

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<sup>3</sup> These concerns include nutrient management concerns for the dairy industry, identified as a major concern in Washington State, and could include odor violations or other regulatory issues for the composting industry.

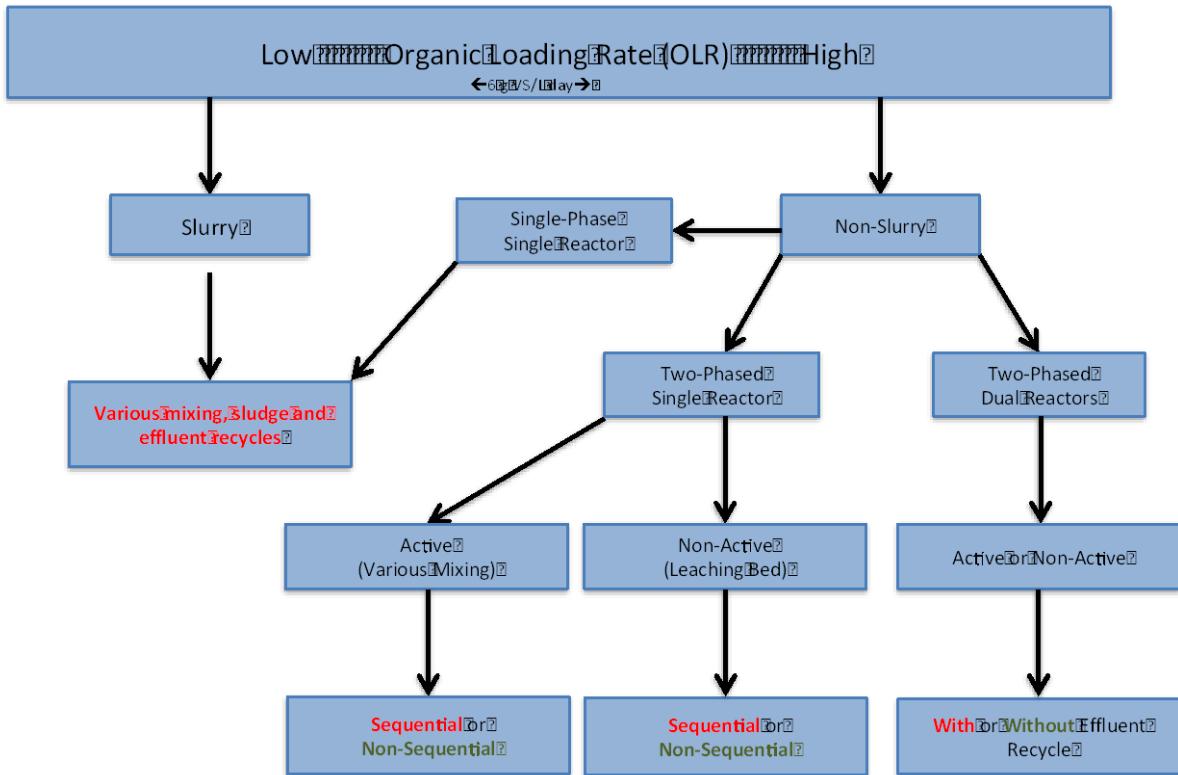


Figure 7. Summary of various OFMSW AD technology approaches.

Figure 7 separates technology into either slurry-based or non-slurry based approaches. Slurry-based approaches actively dilute the solids waste stream with liquid to attain a lower TS level suitable for active mixing. This allows for process homogeneity and presumably improved kinetics, microbial populations, and substrate interactions. Liquid dilution can occur through any combination of water addition, effluent recycle and/or co-digestion. Of these three options, the latter two use less water and generally improve attainment of desirable nutrients, organics, alkalinity and/or economics. Typical slurry systems include common municipal and manure fed systems such as complete mix and/or plug-flow or combinations thereof. At higher organic loading rates (OLR) (rates above 6 g VS/L day by Hartmann and Ahring, 2005), non-slurry digesters begin to out-perform slurry digester in terms of several metrics, such as biogas yield and productivity. This presumably occurs because homogeneous mixing and enhanced degradation kinetics result in rapid product inhibition that cannot be controlled, leading to either depressed performance or complete failure of the digester.

In contrast, non-slurry based approaches refers to processes that do not actively dilute the solids, but instead maintain either a solids bed or a thick paste. This bed or paste is either not actively mixed or, if mixed, is mixed primarily for material transfer rather than for homogeneity. Considerable amounts of liquid may still be used by some of these systems. During early years of OFMSW digestion, slurry systems predominated, presumably because of familiarity with common complete mix designs. However, there has recently been more rapid adoption of non-slurry systems, resulting in a sharing of market percentage between the two types (Figure 8). Bolloniani et al. (2006) state the benefits of reduced reactor volume and wastewater production as primary

reasons for increased market adoption of non-slurry systems, although Hartmann and Ahring (2006) point to concerns with material handling which limits adoption under some conditions.

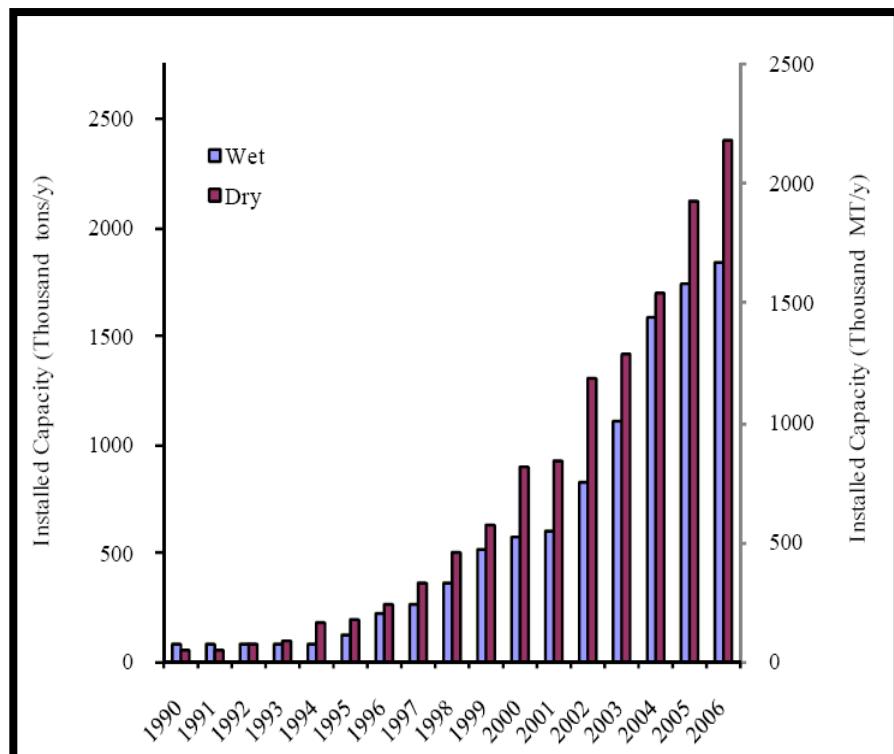


Figure 8. Trends in Slurry (Wet) and Non-Slurry (Dry) AD of OFMSW  
(Hartmann and Ahring, 2006).

## Types of non-slurry systems

### Single phase, single reactor

Several key commercial non-slurry systems as depicted in Figure 9 digest OFMSW with a single phase, single reactor approach. In practical terms they are almost slurry reactors in that they include enough liquid to accomplish an intermediate level of mixing, though much less liquid and higher levels of TS than slurry systems. Mixing promotes a more homogenous mixture, though it primarily is done in order to transport and transfer material.

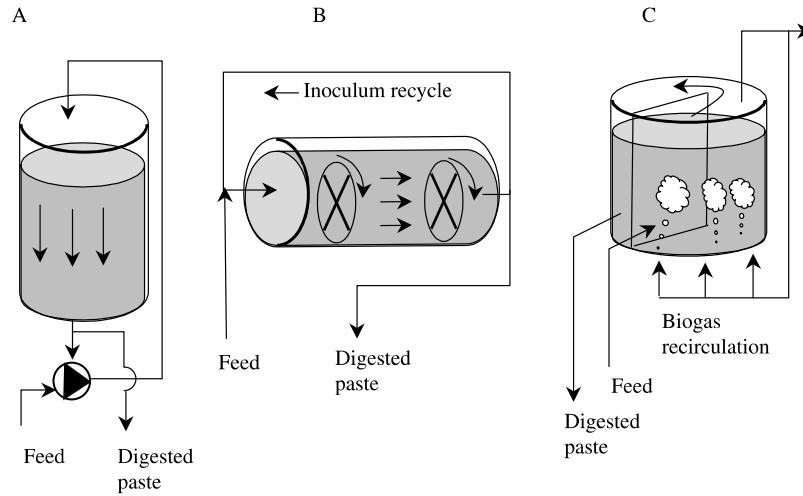


Figure 9. Varying digester designs in single phase, single reactor approach (A is Dranco; B is Kompogas, C is Valorga, from Lissens et al., 2001).

When digesting particularly high strength and highly biodegradable material such as food scraps and/or green waste, it is important to control the degradation kinetics to avoid producing acidic inhibitors which can sour or depress the entire system. To do this, high BOD digesters typically use extensive sludge or effluent recycle, as high as 60% by mass in some cases (Guendouz et al., 2008). This recycle is presumably rich in acclimated bacteria, high in methanogens, and concentrated in high alkalinity (acid buffering capacity), and bulking material. This recycle flow moderates digester acidity and buffers pH. It also elevates pumping/mixing costs and reduces fresh solids flow through (Zaher et al., 2009) though reactor volumes may be reduced as compared to slurry systems. Since mixing at such high solids content is unable to completely achieve homogeneity, zones may form distinct phases that accentuate certain processes and work against overall microbial co-operation, which are vital to stable operation, particularly with organically rich feeds like food scraps (Hartmann and Ahring, 2005). Such a pseudo-phased system can be prone to instability. It can also experience material handling challenges not just because of the complexities of moving and pumping pastes but also because OFMSW is often poorly pre-sorted, including plastics, metals, and other debris that can plug pumping systems (Hartmann and Ahring, 2005).

### **Two-phase dual reactor**

In comparison to single-phase systems, two-phase systems involve a Phase I or hydrolysis/acid forming process and a Phase II or methane forming process. Through reactor configuration and operating conditions, the dual reactor separates hydrolysis/acid formation from methane production with the goal to optimize bacterial community populations and chemical products in two distinct reactors. Phased digestion is of particular interest to practitioners of high strength waste digestion as the highly biodegradable material is already prone to accelerated hydrolysis/acidification and subsequent intermediary chemical acid production that is inhibitory to methane-forming populations. Process designers simply allow this predilection to occur to its full extent, thereby developing two distinct biological processes that in this case are physically separated by two distinct reactors. Advantages of a two-phase system are commonly understood to be that: (1) microbial communities are optimized for respective and subsequent biological steps, thus accelerating digestion; (2) digester stability is improved by system design and operation controls

(as opposed to biological limits such as pH and ammonia inhibition); and (3) by design, distinct microbial populations are built and maintained, in particular methane-forming bacteria that grow slower and are harder to accumulate Cohen et al., 1979).

Veeken and Hamelers (2000) have shown that accumulation of acid products can cause inhibition to not only methane-formers but also to those organisms responsible for hydrolysis and acidification. This is extremely problematic to proponents of two-phased systems as the first phase which originally had been proposed for development of a distinct and optimized process for production of acids may not be optimized for production of acids. This may result in digester inhibition from acidity if levels become too high and/or pH drops too low.

Other researchers have further detailed the inhibitory effects of these intermediate metabolites showing that un-ionized volatile fatty acids (VFA) were key compounds impacting both acidogenesis (Garcia et al., 1991) and hydrolysis of particulates (Llabres-Luengo and Mata-Alvaraz, 1988). In addition, excessive hydrogen formation during hydrolysis also suppresses acid forming reactions (Fox and Pohland, 1994). This has been seen in practice by researchers who observed run-away acidification inhibiting all biological processes (Veeken and Hamelers, 2000; Jiang et al., 2005; Chen et al., 2007). Given the aforementioned process concerns, required modification, and complexities of operation, many in the field have concluded that phased digestion proponents have not effectively proven their case to merit commercialization (De Baere, 2000). As a result, single-phase treatment is still the predominant AD treatment applied at full-scale for OFMSW (Figure 10). To address this concern, our research has modified the phased approach, achieving a controlled, phased system.

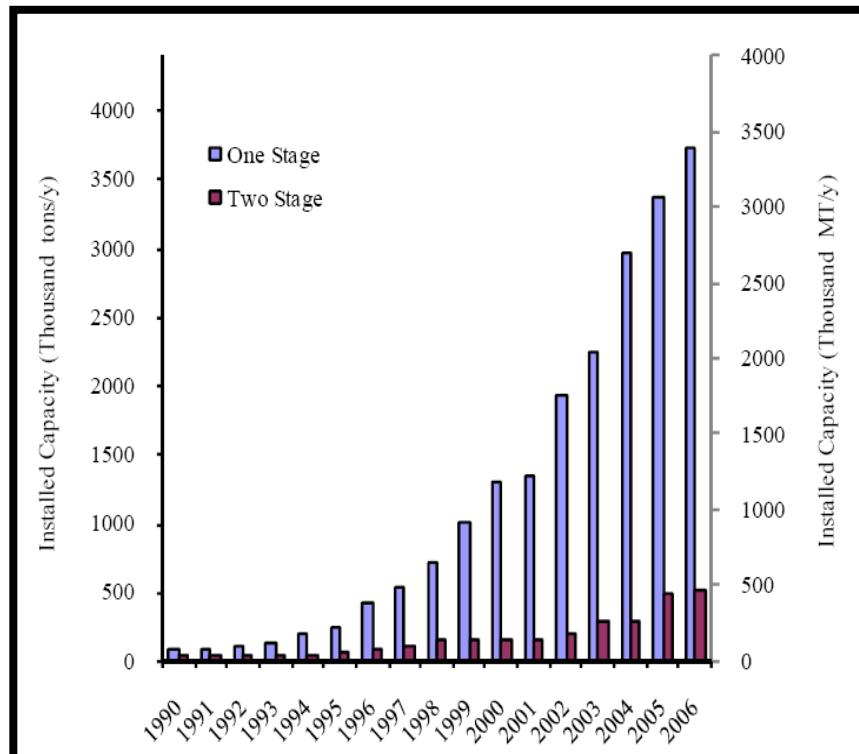


Figure 10. Commercial application of single-phase and two-phase systems by year as against total installed capacity (Hartmann and Ahring, 2006).

## Batch leaching bed reactors

Numerous researchers (Ghosh, 1985; Chynoweth et al., 1992; Vieitez and Ghosh, 1999; Veeken and Hamelers, 2000) have proposed a simple and elegant solution to runaway hydrolysis/acidification in a phased system (Figure 11). This approach is to use a wastewater treatment standard—leaching bed reactors (LBR), sometimes called percolating AD. LBR are single-stage column reactors operated in batch mode, where liquid evolving from the solid bed (leachate) is collected at the bottom of the column. It is assumed that the leachate can assist in mass transfer and substrate/microbial contact, allowing for acceleration of solid-state liquefaction/hydrolysis as well as acidification for production of VFA and alcohols, while simultaneously removing product inhibitors from the solid-bed so that high rate degradation can be maintained (Dogan et al., 2008). While other more active approaches have been utilized, involving mixing and/or continuous flow (Lee et al., 1999; Pavan et al., 2000), a growing body of research and increasing commercial interest involves the batch LBR approach because of its ability to solve runaway acidification concerns in a relatively simple/inexpensive liquid exchange system. In addition, unlike the single phase reactor systems discussed above, batch loading and unloading of the static bed allows for the presence of contaminants, as only liquids are pumped, not solids (Ghanem et al., 2001; Han et al., 2004; Jiang et al., 2005; Chen et al., 2007; Lehtomaki et al., 2006; Cysneiros et al., 2008). The system thus avoids the need for OFMSW pretreatment.

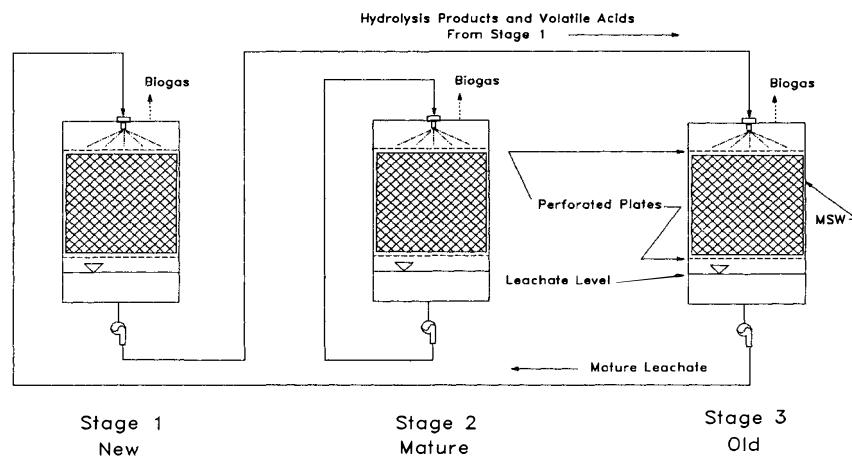


Figure 11. SEBAC system as example of sequential use of single reactor, phased LBR (Chynoweth et al., 1992).

Over time, the various designs involving batch LBR have improved upon overall system stability, and reduced capital and operational inputs.

- *Multiple LBR sequentially fed*—Chynoweth et al. (1992) developed the SEBAC (Sequential Batch Anaerobic Composting) system, which now has been adopted in various permutations by many researchers and commercial entities. In its most basic form, leachate from a starting batch is fed to a later stage batch LBR so that liquid, with high levels of inhibitory VFA, can be converted to methane via accumulated populations of methanogens that reside within the later sequencing LBR. Meanwhile, leachate from this SBR, now high in pH and alkalinity due to removal of VFA and production of buffer system species is sent to the starting batch in the LBR to control hydrolysis/acidification. With proper movement of the leachates system

equilibrium is achieved and substrate degradation proceeds efficiently. In essence the system uses multiple single reactor, phased-systems and does not utilize a dedicated central methane-producing reactor. Instead, each LBR has a lifetime that moves from primarily acidogenic to primarily methanogenic.

- *Multiple LBR non-sequentially fed, utilizing dedicated methane-producing reactor*—Hofenk et al. (1984) and Ghosh (1985) and subsequently many other researchers and commercial entities have explored this two-phase, dual reactor approach where a batch LBR is operated throughout its lifetime as a hydrolysis/acidification reactor whose leachate is sent to a dedicated methane-producing reactor, preferably a high rate granular reactor such as an upflow anaerobic sludge blanket (UASB) digester (Lettinga, 1995). High pH and high alkalinity effluent from the methane-producing reactor, presumably containing some bacteria, is then sent back to the LBR for control of acidification. Importantly, while the LBR is controlled for run-away acidification, it is still operated at a lower pH to induce phased production of acids without production of methane. Hartmann and Ahring (2005) concluded that if phased operation were to be used, use of dual reactors under this regime would be preferred over sequential operation of single reactors. Little experimental data comparing the two was given supporting this conclusion. Conversely, Hofenk et al. (1984) reported process control difficulties with LBR percolation and deterioration of USAB granules due to leachate inhibitors.
- *Non-sequentially fed, single reactor LBR*—Hedge and Pullammanappallil (2007) reported a slight but potentially important modification to LBR process control. Instead of sequentially operating multiple LBR, they determined that a single LBR once operated to stable completion (perhaps using outside controls) could be re-started with new solids using only the cumulative leachate from the first batch, requiring no additional seed or input of external leachate/liquid. While their stable operation was for a particular solid waste mixed with less biodegradable wood chips, they did show the possibility for reducing process complexity, requiring only one phase shifting reactor, without sequencing or a dedicated methane-producing reactor. It should also be noted that significant inhibition to the system was seen after three runs with the effluent due to accumulation of non-VFA inhibitors, namely ammonia.

## Conclusion

In summary, the following general conclusions can be made regarding the various available technology processes for digestion of OFMSW.

- Slurry digesters will undoubtedly remain a major fraction of the market for at least the next several years, due to their simplicity of design and years of practical and operational experience. However, ever-increasing economic pressures related to water demand, footprint and volumetric and material economics may dictate a declining market fraction in the future.
- Continuous flow non-slurry digesters using a non-phased approach have received considerable attention and currently represent a large fraction of the world market. However, interest in the US is more focused on batch, sequential and/or two-reactor systems, perhaps because of the smooth integration with US composting facility infrastructure and operational methodologies. In order to grow the AD market for OFMSW in the US and attain a higher market share, both systems will need continual process

improvements that increase stability and reliability and decrease capital and operational costs.

- While adding complexity and therefore the potential for issues regarding stability, economic cost and operational complexity, two reactor systems operated in batch mode do offer unique co-digestion, feedstock receiving options, and nutrient recovery scenarios. Combined with the environmental benefits these systems could provide, this indicates the potential for commercial interest in the future, assuming that economics can pencil out for projects.

# **Section 3: Floor scale evaluation of multi-reactor, liquid recycle high solids anaerobic digestion system**

## **Background**

In the previous section, it was seen that technological refinements continue to be made for digestion of OFMSW. While slurry digesters using traditional complete-mix and mixed plug-flow designs are still being built, a general movement towards a non-slurry approach is ongoing due to perceived gains in volumetric productivity and lower water usage. Within this sector of non-slurry digesters, processes utilizing batch leaching-bed reactors (LBR) have become commonplace, thanks to the fact that OFMSW pretreatment/contamination removal needs are less stringent than for other approaches (from an AD operations perspective, although retained contaminants are still in need of separation/disposal for quality compost product development), and the fact that they integrate well within existing compost yard facilities and equipment. The use of liquid flow and pumping (as opposed to solids and paste flow and handling required by continuous-flow, actively mixed, non-leaching bed approaches) eases operational demands and enhances stability. Building on these advantages, the goal of this research project was to identify a new LBR concept, slightly altered from that of existing LBR designs. The system was built and evaluated at floor-scale to determine whether the concept was worth future commercial consideration.

## **New leaching bed reactor concept**

Figure 12 is a schematic drawing describing the flow patterns and approach of the design. Briefly, since large compost facilities will be receiving food scraps and green waste on a daily basis, multiple batch reactors will be required to suitably treat the incoming daily flow. While such multiple reactors can cost more than continuous flow systems, this multiple LBR approach could be cost effective on a project basis because of its modular design, smooth integration with compost operations, and removal of a need for dedicated negative-pressure storage buildings. When a LBR is loaded with food/green material, it is mixed with residual digested solids and leachate from a LBR that has completed its digestion cycle. Immediately after loading and sealing of the new LBR, a high flow rate of pre-digested, ammonia-stripped, high alkalinity UASB effluent will be leached through the solids bed, inducing accelerated hydrolysis and acidification that controls pH near or above 6 and thus avoids or drastically minimizes development of an acidified or phased reactor.

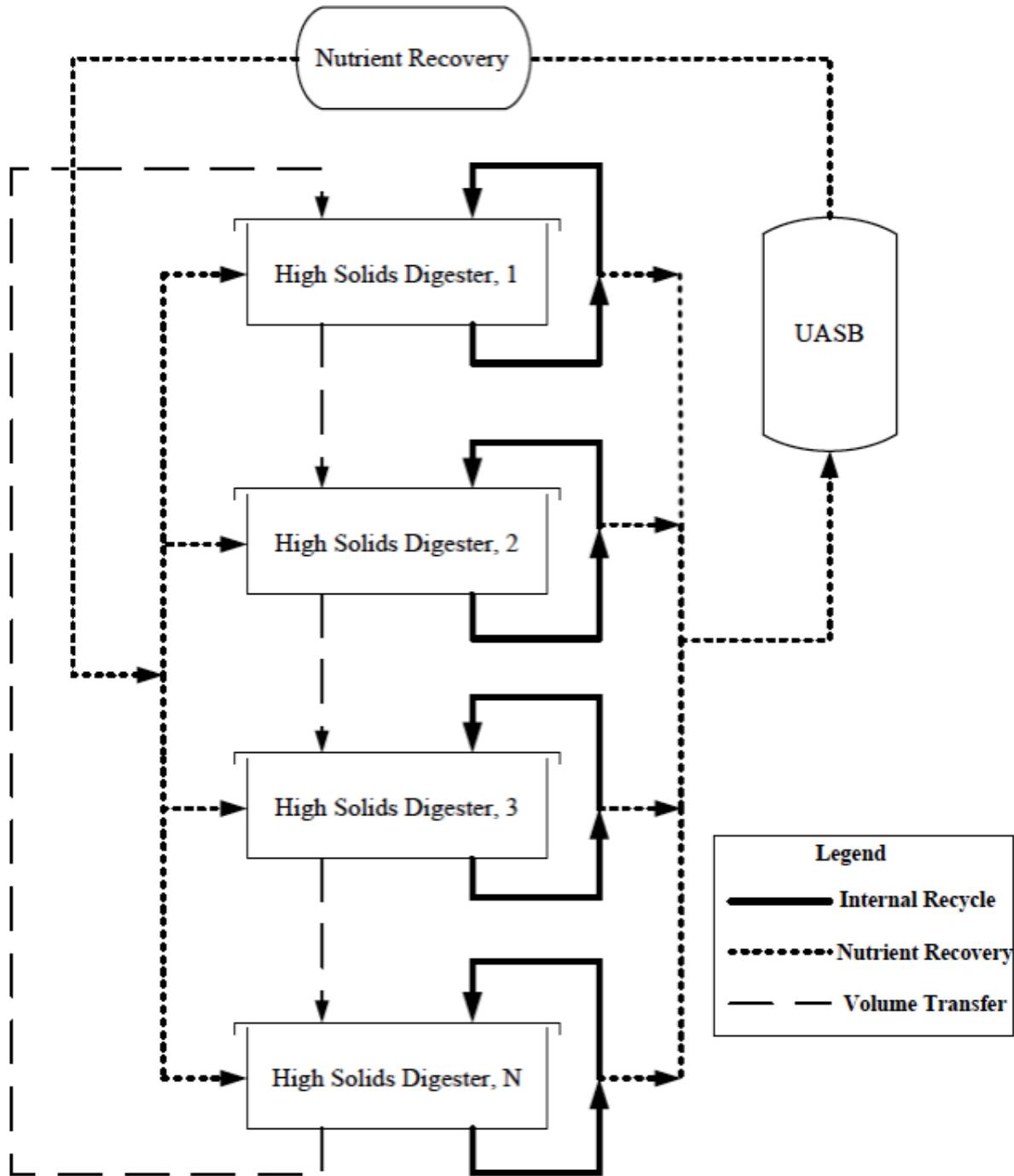


Figure 12. Schematic, flow pattern and design approach of new LBR system for treatment of food scrap and green waste composed OFMSW within a compost yard scenario.

While flow will be high during this limited time period, generating associated liquid pumping cost, concern for out-of-control acidification will be minimized while hydrolysis/acidification can occur in a *non-phased* but optimal environment. Additionally, VFA concentrations and pH within the leachate will be ideally suited for the downstream UASB reactor. Of equal importance, the short duration high-rate flow will allow for near-complete stripping of ammonia from the aged leachate

that was placed in the new reactor, allowing the system to run effectively for the rest of the individual LBR cycle as well as throughout numerous cycles. This drastically reduces the need for fresh water to the system. Once optimal leaching has been achieved and run-away acidification and pH depression has been fully avoided, the LBR switches to an exclusively internal leachate recycle loop, using its own leachate to spur mass transfer, degradation kinetics, and accumulate alkalinity for additional stability. During this process, methanogens, either active since the original seeding or seeded from the bacteria present in the recycled, digested UASB initial effluent will rapidly convert VFA—generating the desired renewable energy while maintaining optimal low levels of acidic inhibitors and preparing the leachate solution for use within the next LBR.

## Theory

As described in Section 2, the bulk of this system builds upon previously described and implemented work (Hofenk et al., 1984; ten Brummeler, 2000; Hedge and Pullammanappallil, 2007). The design incorporates subtle process changes and integrations that are based on review of literature and available science. Design modifications include the following:

- *Short duration, high flow rate leaching during initial stage*—Veeken and Hamelers, (2000) showed that almost immediately upon loading, accelerated hydrolysis and acidification will occur within an LBR, potentially depressing the pH through VFA production to the point of inhibition of methanogenesis as well as acidification and hydrolysis. Veeken and Hamelers (2000) originally proposed that initial acidification and pH depression is most effectively prevented by an LBR process that reduces initial leachate flow. However, we hypothesize that it is more effectively prevented through an initial high flow of leachate, as has already been indirectly shown by Lu et al. (2008).
- *Non-sequential LBR and use of volumetrically low levels of seed*—While some processes (i.e. SEBAC) utilize a sequential approach to minimize or remove seeding requirements (Chynoweth et al., 1992), this process can be complex. Our non-sequential LBR approach relies on high levels of seed, returning as much as 65% of originally digested solids back to the new LBR for acidification and pH control (ten Brummeler, 1990). This approach is supported by Hedge and Pullammanappallil (2007), who have shown that finished effluent from an initial LBR can serve as a sole source of mass transport and chemical/biological species media for a second LBR, requiring no solids seeding. A non-sequential approach between LBR is developed here but with the use of small volumetric loadings of seed. In short, the process builds upon the seeding capabilities of a two reactor system (LBR/UASB) while minimizing both the complex sequential LBR cycling and the need for excessive waste solid/seed return.
- *Non-phased, two reactor approach*—As described in Section 2, researchers still find sufficient data to conclude that phased digestion leads to process improvements (De Baere, 2000; Hartmann and Ahring, 2006). Numerous sequential and non-sequential LBR designs still purposefully or with resignation, allow their LBR to lower to pH below 5 and even maintain their operation for long-periods of time at this depressed but controlled pH, only later upon equilibrium achieving a higher pH and a working methane population (Chynoweth et al., 1992; Hedge and Pullammanappallil (2007). Two reactor proponents (LBR/UASB) also primarily operate their LBR as an acidifying, phased reactor, producing methane almost exclusively within the UASB (Hofenk et al., 1984; Lehtomaki et al., 2006). While Wu et al. (2005) showed that both hydrolytic and acidogenic bacteria can tolerate wide fluctuations in environmental

conditions and lose little to no activity within a pH range of 3-7, others have shown that optimal performance occurs at more specific ranges, with acidogens targeted for a pH of between 4-6.5 (Speece, 1996) and hydrolytic bacteria at pH of 6.0-6.5 (Arntz et al., 1985) or even 7.0 (Cysneiros et al., 2008). Eastmann and Ferguson (1981) demonstrated that acidogenic processes are mainly regulated by the hydrolytic step and therefore acidifying kinetics is best determined by hydrolysis. From this they inferred that while acidifiers might perform optimally over a wide range of pH, improved performance will be seen when operating at a pH of 6.5 or above where hydrolysis is preferred and acidogenesis is still active. Of equal importance is that this higher pH reduces the amount of time during which H<sub>2</sub>, an acidogenic inhibitor, accumulates (Fox and Pohland, 1994). Based on this knowledge, our system aims to operate the LBR at near neutral pH, acting as a single-phase reactor.

- *Phased insertion of bacterial populations*— The LBR operates with an initial external recycle loop and a later internal loop. This allows microbial populations to be optimized for the degradation process occurring. The initial loading of the LBR involves two sources of bacterial/archae population, the mature leachate and solids seed from a previous LBR that contains a full consortium of populations, in particular hydrolytic and acidogenic. Thus, during the initial start-up, there will be a large concentration of necessary hydrolytic and acidogenic bacteria. Later, these populations are supplemented with methanogens (influent or bacteria) as the LBR continues to operate at stable methane-producing levels.
- *Nutrient recovery for removal of inhibitors, water savings and additional product revenue*— Previous work within the area of animal manure AD research and development has indicated that project adoption can be spurred when the project generates multiple revenue streams, and when it provides a mechanism to resolve important environmental/regulatory concerns facing adopters, namely nutrient loading and release in the case of dairy AD facilities (Frear et al., 2011). A similar strategy may be important for HSAD, given that large compost facilities are regulated on nutrient releases from air and water. Unfortunately few AD operations, either farm-based or municipal, currently practice active nutrient recovery. In this system, nutrient recovery will be incorporated within the initial LBR/UASB feed loop. This maximizes the potential for recovery and removal of inhibitory ammonium from the LBR, while limiting flow through to the nutrient recovery system.

## Goal and objectives

The overriding goal of this research project was to validate proof of concept of the entire integrated process as well as individual unit operations and associated hypotheses, using a set of floor-scale LBR reactors. Specific objectives included:

- Determine theoretical methane potential, energy production and nutrient recovery from test feedstock, a mixture of cafeteria food scraps and green waste simulating the type of feedstock that would typically be available at a compost facility.
- Test the batch, internal recycle of LBR for phased digestion in a single reactor to determine effectiveness and whether or not the process requires an external loop leaching to operate effectively.

- Optimize and develop a model for high-rate, short-duration external leaching of LBR for initial pH and acid production control. Factors of leachate rate and alkalinity concentration will be investigated against pH, VFA, SCOD, biogas, methane content, and hydrogen content indicators.
- Assess the full, proposed floor-scale system for yields and productivities.

## Section 4: Mixed food scrap and green waste characterization, specific and ultimate methane potential and nutrient recovery capability

### Background

Hartmann and Ahring (2006) summarized a wealth of performance data from OFMSW AD in Europe (Figure 13). The majority of commercial digesters operating on various fractions of OFMSW have volumetric biogas production rates somewhere between  $1.5 \text{ m}^3/\text{m}^3\text{/day}$  and  $3.5 \text{ m}^3/\text{m}^3\text{/day}$  (biogas production curves). Most of the digesters that are performing at or above this range operate at high OLR, above the previously described slurry/non-slurry threshold of  $6 \text{ kg VS/m}^3\text{/day}$  and closer to  $10-15 \text{ kg VS/m}^3\text{/day}$ . Within this population, several systems are operated on a combination of OFMSW, food scraps, and yard waste, producing an average of roughly  $0.35-0.55 \text{ m}^3 \text{ biogas/kg VS}$  in feedstock (data from Hartmann and Ahring, 2006, not shown here). When converted to  $\text{CH}_4$  using an assumed average methane content, these approximate numbers can be compared to the particular feedstock being utilized in this study and its calculation for theoretical and specific (experimental) methane potential ( $B_u$  and  $B_o$  respectively;  $\text{m}^3 \text{ CH}_4/\text{kg VS}$ ) as well as volumetric methane production rate ( $\text{m}^3 \text{ CH}_4/\text{m}^3\text{/day}$ ).

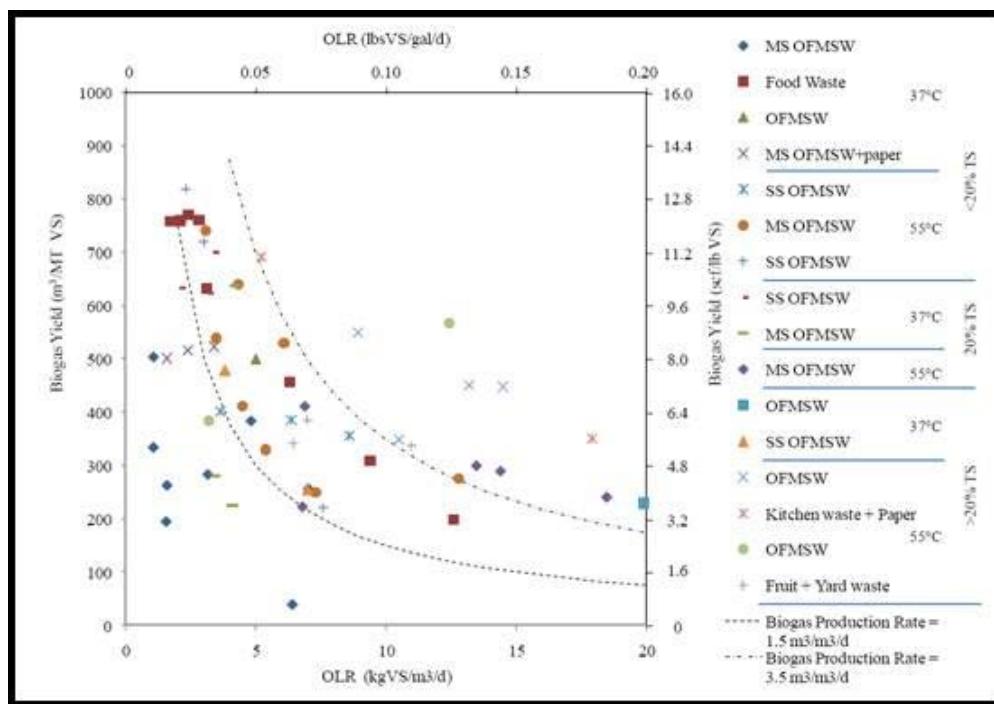


Figure 13. Overview of methane specific biogas productivities at commercial facilities, plotted against OLR, waste mixture, and volumetric biogas production (Hartmann and Ahring 2006).

### Synthesis of solid mixture

A representative solid waste mixture that would be conceivably available at a US compost facility was generated using Washington State University (WSU) cafeteria food scraps and a mixture of freshly cut green grass and barnyard straw. A compost yard front loader was used to combine

approximately 70% food scraps and 30% mixture green grass and straw, and deliver the mixture to a compost yard chopper/mixer normally servicing the yard for preparation of compost piles. Waste sources were contaminated with a variety of materials again representative of what might be delivered to a compost yard, including biodegradable bags, plastic utensils, seeds, etc., and all material was placed in the mixer/shredder. The mixture was chopped/agitated for 15 minutes and then delivered via belt press to a number of clean 5 gallon buckets, which were then sealed and stored in a walk-in cooler at 4 °C until it was used for experimentation. Figure 14 is an image of one of the full 5 gallon buckets.



Figure 14. Example of one of the solid waste mixture buckets collected, stored and utilized during project research.

### **Characterization of solid mixture and other digester inputs**

Table 3 is a summary of the physical and chemical characteristics of the food scrap and green waste mixture that was developed for this study. Characterization was also completed for anaerobic granules (accumulated biomass within an operating starch-water UASB reactor, Penford Foods, Kennewick, WA) that were used as inoculum in the studies, as well as UASB effluent that was utilized as external leachate. As can be seen from the table, the food waste mixture is of a high solids content, roughly 30.3% TS assuming a density equal to water. The material has a very high VS/TS ratio (95%), indirectly indicating a high organic, but not necessarily biodegradable organic content. Less than 3.5% of solids are inert, fixed or ash in content. Additional measurements point to a large COD and high TKN, again indicating a high degree of organic material capable of chemical if not biological conversion but containing a significant amount of nitrogen, primarily in

the form of organic N (proteins) but some inorganic ammonia. Total phosphorus is quite low while the rapid diffusion into water showed a quick conversion to low pH, due mostly to the rather high levels of VFA and little to no alkalinity. Additional proximate analysis of the food waste mixture showed that it was composed (wet basis) of 6.06% lipids, 5.55% protein, 0.407% VFA, 1.48% ash, and 15.99% non-VFA carbohydrates.

Table 3. Influent parameters of the food scrap and green waste mixture used in this study.

Parameter (g/L)	Food Waste Mixture <sup>a</sup>	Anaerobic Granules <sup>a</sup>	UASB Effluent <sup>a</sup>
TS	303 ± 1	87.0 ± 0.9	4.63 ± 2.60
VS	288 ± 1	72.1 ± 0.5	0.16 ± 0.40
FS	10.55 ± 0.07	15 ± 1	7.4 ± 0.8
COD	370.7 ± 1.76	151.1 ± 0.45	0.865 ± 0.013
VFA	4.07 ± 0.45	N.D.	N.D.
TKN	8.44 ± 0.29	7.97 ± 0.06	0.0811 ± 0.022
TAN	1.23 ± 0.01	N.D.	0.028 ± 0.003
TP	1.04 ± 0.26	0.363 ± 0.005	0.044 ± 0.007
pH	3.82 <sup>b</sup>	7.59	8.3 ± 0.3
Alkalinity	3.48 ± 0.05	3.03 ± 0.04	1.5 ± 0.50

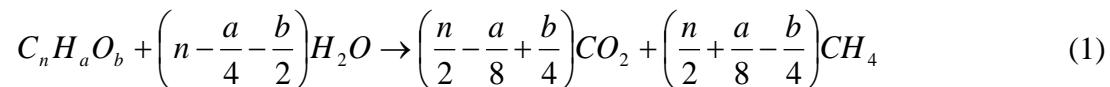
<sup>a</sup> Data is the average of (n=6) with mean standard deviations at α=0.05

<sup>b</sup> Food mixture was placed in distilled water, gently mixed and then recorded for pH

UASB effluent was obtained from a UASB reactor in Kennewick WA digesting waste potato starch water (Penford Foods). This effluent was chosen for its proximity to WSU as well as its high degree of digestion (low VFA and COD), relatively high pH, moderate to low levels of alkalinity and extremely low levels of nitrogen. Combined, these characteristics made the effluent an excellent candidate for mimicking ammonia stripped UASB effluent that could easily be dosed to be representative of additional treatment levels of alkalinity.

## Biogas and methane productivity calculations

The Bushwell formula has been used to calculate theoretical maximum biogas and methane production, or  $B_u$  values, given experimental data for composition of particular wastewaters, including municipal and agricultural wastewaters entering anaerobic digesters (Moller et al., 2004). This value provides an upper limit for actual specific methane productivity that can guide experimentation and which can be used to interpret the experimental data obtained from specific systems. The Bushwell formula is as follows:



using the following assumptions:

- VS<sub>lipid</sub> (C<sub>57</sub>H<sub>104</sub>O<sub>6</sub>); VS<sub>protein</sub> (C<sub>5</sub>H<sub>7</sub>O<sub>2</sub>N); VS<sub>carb</sub> (C<sub>6</sub>H<sub>10</sub>O<sub>5</sub>); and VS<sub>VFA</sub> (C<sub>2</sub>H<sub>4</sub>O<sub>2</sub>).

- Specific methane productivity values for VFA, carbohydrates, proteins, and lipids are 370, 415, 496 and 1014 m<sup>3</sup> CH<sub>4</sub> /kg VS, respectively.
- VS total = 288.0 kg/m<sup>3</sup> with respective VS fractions being 19.82%, 57.10%, 21.64%, and 1.45% for protein, carbohydrate, lipid and VFA, respectively.

Calculation of the theoretical B<sub>u</sub> values for the solid mixture yields a value of 0.56 m<sup>3</sup> CH<sub>4</sub> /kg VS. This is (and should be) above the actual specific methane productivities (B<sub>o</sub>) reported by Hartmann and Ahring (2006) for mixtures of food scraps and/or green waste (~0.3-0.4 m<sup>3</sup> CH<sub>4</sub> /kg VS). Thus 0.56 m<sup>3</sup> CH<sub>4</sub>/kg VS can be viewed as an upper limit for actual specific methane productivity. Meanwhile a value more in the range of 0.34 CH<sub>4</sub>/kg VS is to be experienced in this floor scale digestion project, with actual target range to be determined upon experimentation.

Since the VS of the material is 288 kg VS/m<sup>3</sup> and a typical range of digestion time for LBR systems is 21-42 days, it is conceivable to assume that this LBR system would, at an average solids retention time of 28 days, have an organic loading rate (OLR) of roughly 10 kg VS/m<sup>3</sup>/day. This again places the system well within the range of typical OLR for non-slurry or high solids reactors. With this OLR, the system should achieve an effective volumetric loading rate of between 1-4 m<sup>3</sup> biogas/m<sup>3</sup> day, consistent with Hartmann and Ahring (2006).

After inclusion of nitrogen data at an assumed nutrient recovery rate of 80% of total produced ammonia, or 50% of influent TKN (Frear et al., 2011), Table 4 summarizes expected minimum performance results if the system being verified can work up to existing commercial digester capabilities for the feedstock being digested.

Table 4. Target range for biogas values, methane values, and overall performance outcomes for experimental reactor system.

	Target Biogas Value	Target Methane Value (~65%)
Theoretical Productivity (m <sup>3</sup> /kg VS d)	0.86	0.56
Specific Productivity (m <sup>3</sup> /kg VS in)	0.59	0.34
Volumetric Production Rate (m <sup>3</sup> /m <sup>3</sup> /day)	1.5-6	1-4
Target Value		
VS Destruction (%)	60	
Organic Loading Rate (kg VS/m <sup>3</sup> /day)	10	
Nitrogen Product (kg/MT dry solids)	14.7	

## Section 5: Experimental methods

### Experimental system

Figure 15 depicts the experimental system comprised of three parallel anaerobic reactor units capable of individual experimental operation. Each unit consisted of a 37 liter leaching bed reactor (LBR) containment tank, 10 L buffer tank, 35 L mixed, effluent storage tank, and supporting gas collection tipping buckets, peristaltic feed pumps, feed timers, pH/temperature meters and data loggers. All tanks were maintained at 38 °C using internal hot water heat exchangers fed by an electric water heater and water-recycle pump. Heat losses and corresponding energy demand were minimized through the use of wrap insulation.



Figure 15. HSAD Experimental system.

Each LBR containment tank served as a heating vessel, pressurized gas headspace collection site, and physical housing for pre-loaded LBR reactor buckets (Figure 16). Upon immersion in an internal water bath, the containment tanks maintained mesophilic temperature using heat exchange between the internal hot water pipes of the containment tank and the internal water bath.

Headspace gas produced during reaction vented to a calibrated, water-displacement tipping bucket (wet tip gas-meter, TN, USA) attached to a data logger (Onset, MA, USA). Periodic testing of gas quality using evacuated gas collection tubes (Labco Exetainer, UK) was accomplished through

insertion of a syringe into a septum port downstream of a collection valve.



Figure 16. LBR reactor bucket design and operation.

While use of a non-submerged down-flow LBR regime was originally preferred, scaling/design issues required that LBR reactor buckets (Figure 16) be operated as submerged, up-flow leaching bed reactors. Influent entered the bottom of the reactor bucket underneath a 20-mesh screen, passing through a food scraps/green waste feed mixture held in place by another 20-mesh screen, and ultimately exiting out the top of the bucket through a gas-tight, U-tube hosing fitted with a solids clean-out. Placement of the U-tube allowed for a constant liquid level in the reactor bucket above that of the solids bed height, ensuring complete submersion of the organic material.

Two peristaltic pumps (Stenner pumps, IN, USA) were used to accomplish both the internal and external recycling to and from the LBR. One pump drew liquid at a proscribed flow rate from the effluent U-tube, back to the influent side of the reactor bucket, thereby accomplishing an internal LBR recycle. Meanwhile effluent from the mixed liquid storage tank was sent through a second pump at proscribed feed rates to the influent side, thereby completing an external liquid feed to the LBR. Overall liquid volume to the LBR was held constant by allowing for an overflow of excess liquid introduced into the system to leave the U-tube for storage, sampling and down-stream treatment within the buffer tank. Digital pH and temperature meters (Omega Engineering, CT, USA) were placed near the bottom of the U-tube for continuous data logging of effluent pH and temperature. An additional temperature probe was placed inside the solids bed for temperature reading and maintenance of the reactor bed itself. A two-point calibration was performed on the

meters before operation of each experimental run. Cleaning and calibration of peristaltic pumps and gas meters were also completed before each experimental run.

As discussed in Section Five, loading of the LBR reactor buckets was accomplished by preparing a 3.0 kg mixture (2.5 kg food/green waste + 0.5 kg inoculums) of feed held in place within the bucket by two 20-mesh screens, above and below the influent and effluent ports of the liquid leachate. The feed was a mixture of previously prepared and 4 °C stored food scrap/green waste, solid seed either from a commercial UASB digester (Penford Foods, WA, USA) or previous LBR run as inoculums, and UASB effluent. Treatments varied in terms of the external recycle rates to the LBR and the alkalinity of UASB effluent utilized (as-produced effluent or artificially dosed with sodium bicarbonate).

## Experimental design

The experimental design is summarized in Table 5. As a first treatment study, the substrate material was batch digested using internal recycle only, in essence digesting the organic material in the same liquid volume with acid and other metabolites building over time. Following evidence described in previous literature, the expectation was that the system would go sour with decreasing pH and accumulation of inhibitory levels of VFA. After validation of this assumption the LBR reactors were operated in a combined internal (IR) and external recycle (ER) with the ER coming from previously digested UASB effluent. Total volume was held constant, allowing for an overflow of metabolite concentrated liquid and entry of fresh liquid to the system. Treatments within this study were varied by ER rate and level of effluent alkalinity while all other parameters including IR were held constant.

The hypothesis here is that inclusion of fresh liquid containing alkalinity and bacteria could assist the system in resisting pH depression and VFA elevation, maintaining a healthy, operating digester over time. However, it is anticipated that continued entry of ER would ultimately lead to excessive use of liquid and dilution of soluble organics in overflow effluent, creating conditions non-conducive to successive UASB digestion. Results from this series of experiments were used to find an optimal operating condition for a second series of experiments aimed at determining a suitable phased IR and ER approach. In essence the strategy involved operating the digester from the start with a high level of ER, but soon phasing the system to total IR treatment for improved performance in regard to healthy, sustained operation and concentration of soluble organics for subsequent UASB treatment.

Table 5. Experimental plan and treatments.

Experimental Runs	T1	T2	T3
Batch IR (Acc. or Non-Acc. seed)	5 L/day Volumetric Internal Recycle		
Batch IR/ER (Alk (4g/L) and IR (5 L/day)	ER 1x	ER 2x	ER 4x
Batch IR/ER (ER 2x)	Alk 2 g/L	Alk 4 g/L	---
Batch Phased IR/ER (ER only first 5 days)	2x/4 g/L/Non-Acc	2x/4 g/L/Non-Acc	---

## Analytical methods

Analysis of common wastewater parameters was carried out on food scrap/green waste mixtures, granular seed, UASB effluent, LBR bed mixtures before and after digestion, and liquid effluents. All analytical methods for the parameters listed below, including total solids (TS, 2540B) and

volatile solids (VS, 2540E) were conducted according to referenced Standard Methods (APHA, 1998). Chemical oxygen demand (COD) was analyzed with a Hach 45600 COD Analyzer (Loveland, Colorado, USA) (5220D). Alkalinity values were analyzed using a Mettler Toledo T50A Automatic Titrator (Schwerzenbach, Switzerland) (2320B). Protein, Total Kjeldahl Nitrogen (TKN), and Total Ammonia Nitrogen (TAN) were analyzed using a Tecator 2300 Kjeltec Analyzer (Eden Prairie, MN, USA) (4500-NorgB; 4500NH<sub>3</sub>BC). Total phosphorous (TP) was digested and analyzed using an O-I-Analytical FS3000 Flow Injected Analyzer (College Station, TX, USA) (4500PB; 4500PE). Fat content was measured using a Soxhlet apparatus (5520D). Carbohydrate content was calculated from subtraction of known protein, fat, moisture, and ash values.

Volatile fatty acids (VFA) C<sub>2</sub>-C<sub>7</sub> including ethanol and lactic acid were analyzed using a Shimadzu Gas Chromatograph outfitted with an AOC-5000 auto-sampler and FID detector operated at 300°C (GC-2014; Sunnyvale, CA, USA) using a HP-INNOWax column of 30.0 m length, 0.25 mm ID and 0.25 μm film. Start temperature was 70°C, ramped at 15°C per minute to a hold temperature of 230°C for three minutes. Split/split-less injector temperature was 250°C at a split ratio of 40:1 using Helium carrier gas at a total flow rate of 41.3 mL/min. Sample preparation prior to GC analysis included centrifuge at 3000 rpm for 10 minutes at which point 4 mL of 1% formic acid is added to 1 mL of sample supernatant and passed through a 0.22 μm syringe filter for insertion into a 2 mL GC vial. Biogas composition, including CH<sub>4</sub>, CO<sub>2</sub>, and hydrogen sulfide (H<sub>2</sub>S) were analyzed using a Varian GC CP-3800 (Palo Alto, CA, USA) following methods detailed in Wen et al. (2007).

# Section 6: Batch digestion with only internal leaching bed reactor recycle

## Background

As a first step in the experimental study, the mixed food/green waste was subjected to two batch digestions. Each of these batch digestions used only internal recycle within the test leaching bed. The aim of this first set of batch trials was to determine if the waste organic mixture would sour, or whether full reactor failure would occur due to improper process control of the complex anaerobic process (only internal recycle of leachate). These outcomes would be expected given digestion of an organic waste stream that was so rich in organics.

The two batch scenarios tested within the floor-scale leaching beds and described fully in Section Five were designed as follows:

- *Batch with non-acclimated seed and liquid*—food/green waste substrate (10 kg) seeded and mixed with UASB granules (2 kg) and then fully immersed in UASB treated effluent (30 L) operated at an internal recycle rate of 30 L/day or 1 liquid turnover per day.
- *Batch with acclimated seed and liquid*—food/green waste substrate (2.5 kg) seeded and mixed with previously digested biomass (0.5 kg) and then fully immersed in previously digested leaching-bed effluent (5 L) operated at an internal recycle rate of 5 L/day or 1 liquid turnover per day.

## Results and discussion

The results of the two batch runs are summarized in Figure 17 through Figure 23. Figure 17 shows pH over time for the acclimated and non-acclimated runs. Both acclimated (blue) and non-acclimated (green) runs experienced steep drops in pH over about the first 4-5 days. However the non-acclimated run reached a much lower pH of 3.8 during the first four days while the acclimated run leveled off at 4.9 during that same time frame. These depressions in pH point to development of an acidified reactor that is often purposefully maintained for phased digestion, but which is not maintained in this reactor design. While acidification can occur at such pH ranges (Speece 1996), additional evidence shows that optimal hydrolysis and acidification occur at ranges notably higher than either of these depressed levels (Arntz et al., 1985; Eastmann and Ferguson, 1981).<sup>4</sup> Therefore it can be inferred that optimal operation and a focus of this design proposal would be to inhibit the depressed pH and instead operate from the initial stages at a higher controlled pH.

As Hedge and Pullammanappalil (2007) have previously demonstrated, use of acclimated seed and/or previously digested leachate effluent can appreciably affect maintenance of a controlled, higher pH and thereby accentuate operation and performance of the LBR. This phenomenon is readily noted in the sequence of figures. First, as Figure 17 shows, the pH of the run utilizing acclimated media leveled off at a pH at least one full pH unit above the run utilizing non-

<sup>4</sup> While Wu et al. (2005) showed that both hydrolytic and acidogenic bacteria can tolerate wide fluctuations in environmental conditions and lose little to no activity within a pH range of 3-7, others have shown that optimal performance occurs at more specific ranges, with acidogens targeted for a pH of between 4-6.5 (Speece, 1996) and hydrolytic bacteria at pH of 6.0-6.5 (Arntz et al., 1985) or even 7.0 (Cysneiros et al., 2008).

acclimated media. A primary reason for this relates to the fact that the starting leachate of the non-acclimated UASB effluent had much lower alkalinity ( $2 \text{ g/L CaCO}_3$ ) than the digested leachate effluent ( $5.6 \text{ g/L CaCO}_3$ ), showing the importance high alkalinity for buffer pH control.

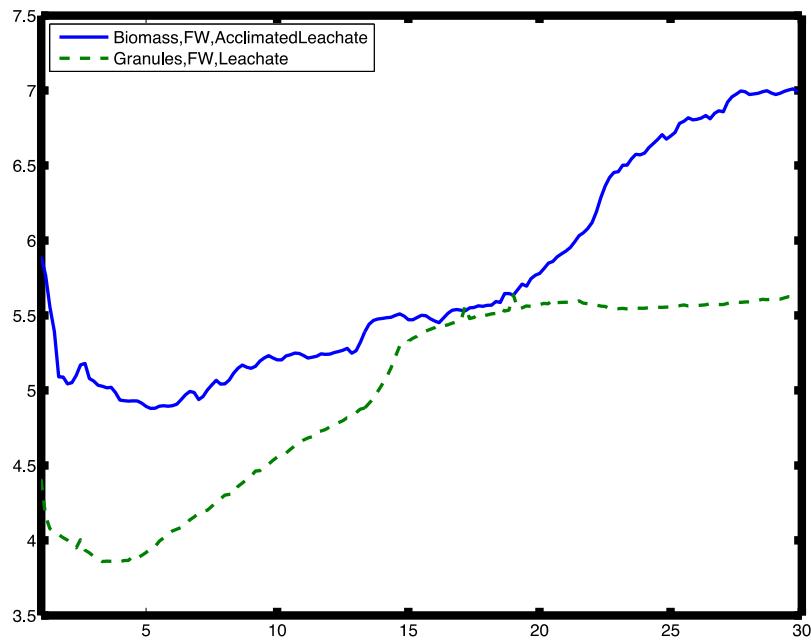


Figure 17. pH against reaction time for non-acclimated and acclimated batch runs.

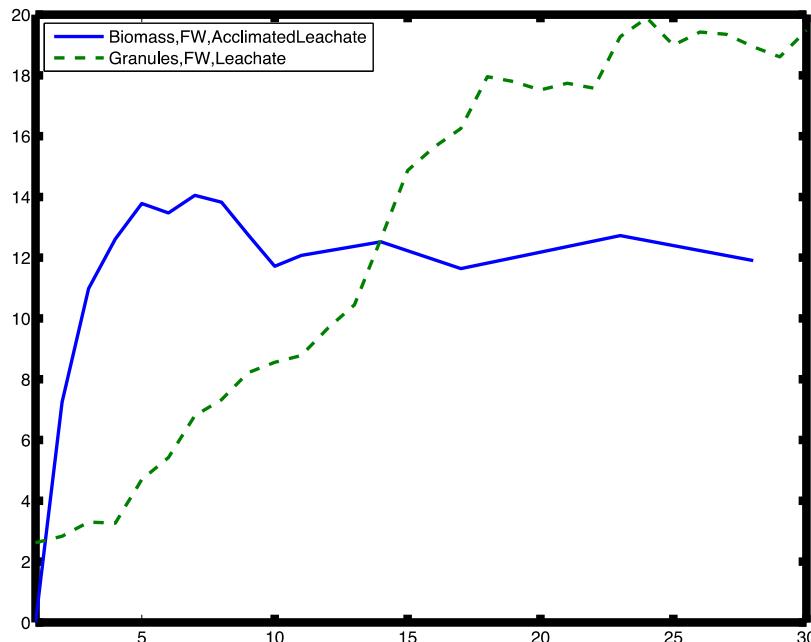


Figure 18. VFA against reaction time for non-acclimated and acclimated batch runs.

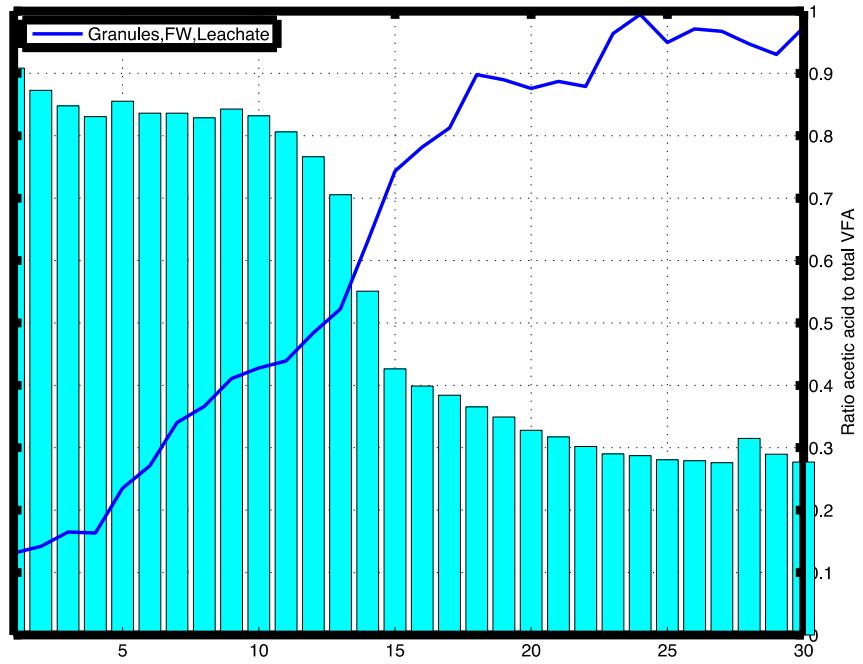


Figure 19. VFA against reaction time with change in percent acetate for non-acclimated batch.

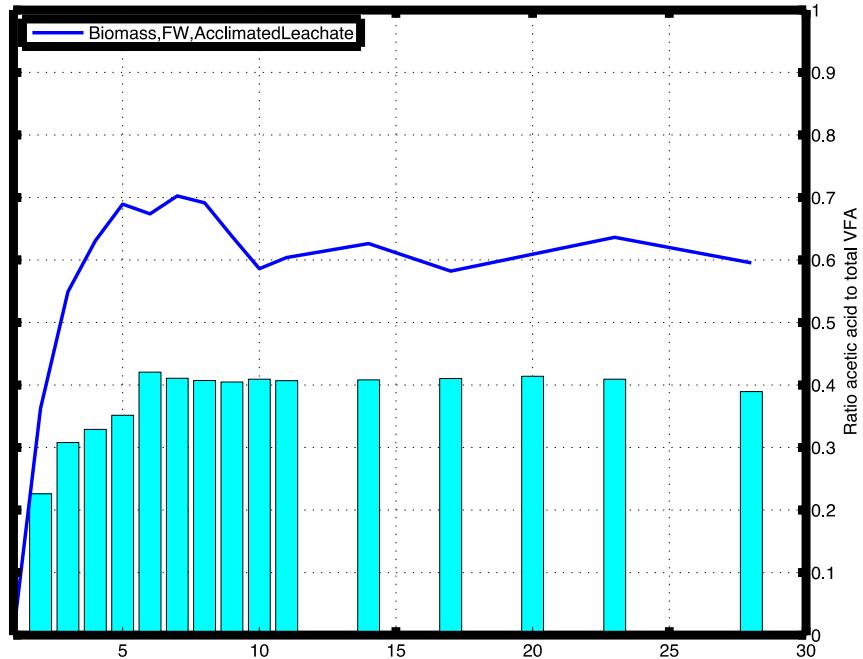


Figure 20. VFA against reaction time with change in percent acetate for acclimated batch.

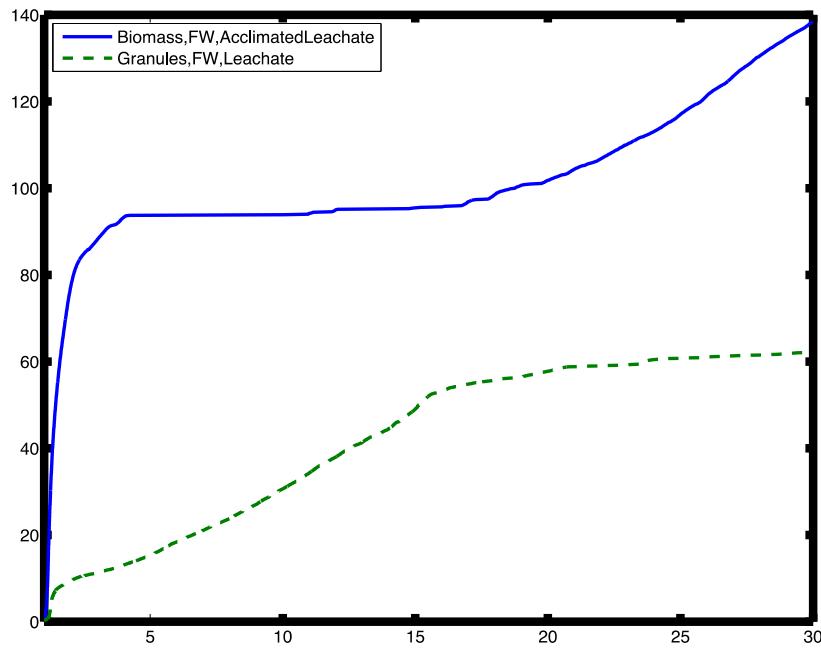


Figure 21. Biogas accumulation against reaction time for non-acclimate and acclimated batch runs.

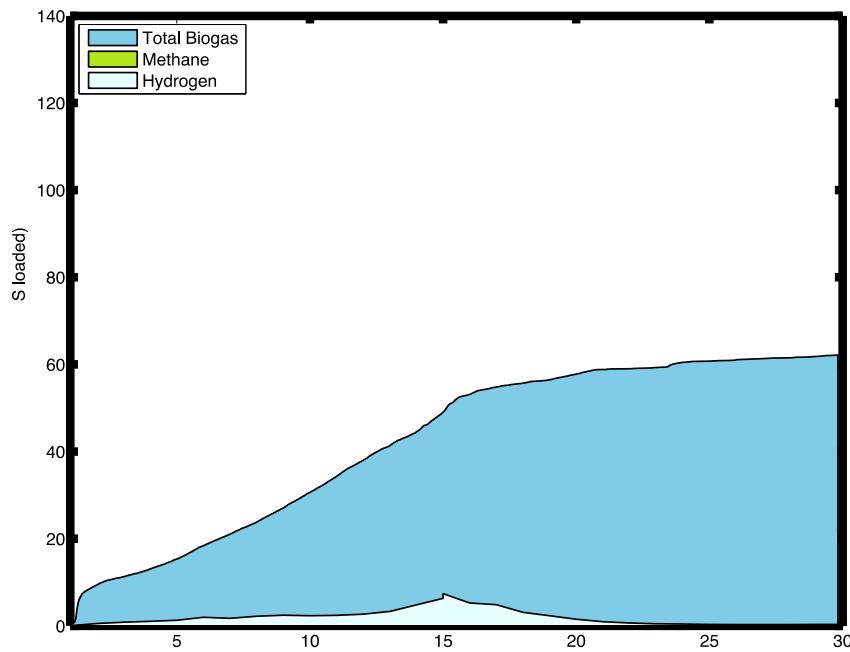


Figure 22. Biogas accumulation against reaction time with changes in biogas constituency for non-acclimated batch run.

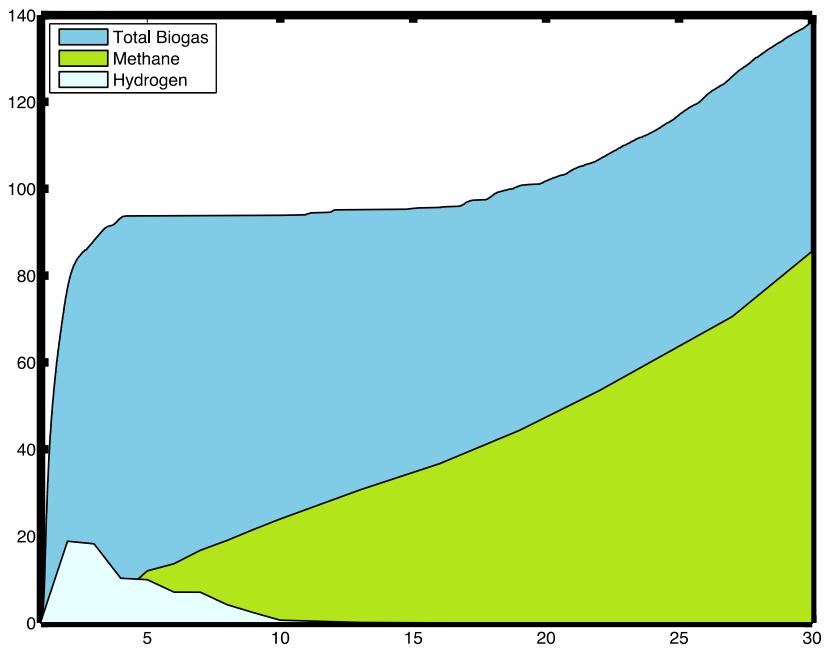


Figure 23. Biogas accumulation against reaction time with changes in biogas constituency for acclimated batch run.

Figure 18 clearly shows the impact of rapid stabilization of pH at a higher and more preferred range, with VFA accumulating much more quickly in the acclimated run compared to the non-acclimated run. By day five, the acclimated run resulted in total VFA concentrations of 14 g/L, a level nearly five times that achieved by the non-acclimated run. With such rapid production of acids, one might worry that a corresponding crash in pH could occur, but (as shown by the rising or relatively stable pH in Figure 17) both the existing alkalinity as well as alkalinity produced during the hydrolysis and acidification steps in fact overrode the tendency for pH decline. Notably, both batch runs resulted in gradual climbs in pH to a level of 5.5 by day fifteen, providing evidence that alkalinity produced from degradation was able to maintain pH despite VFA accumulation.

Figure 19 and Figure 20 show the percentage of acetate in the VFA for the non-acclimated batch and the acclimated batch, respectively. In each of these plots, percentage acetate is plotted alongside the relevant total VFA concentration from Figure 18. The percentage acetate, an easily digested VFA, for the non-acclimated batch is initially quite high (~80%), but later falling to about 30% (Figure 19). In contrast, in the acclimated run, the percentage of acetate is initially quite low (~20%), rising quickly to about 40% (Figure 20).

Figure 21 shows overall biogas production for the acclimated and non-acclimated runs, indicating that at day 15 the two batch runs diverge markedly in their performance, with the acclimated run achieving stability (as shown by ongoing biogas production) while the non-acclimated run sours and eventually fails completely (with cessation of biogas production). The reason for this divergence is the third important point to be considered and is best explained by comparison of Figure 22 and Figure 23, showing biogas accumulation in the non-acclimated run and acclimated

run, respectively, with Figure 17 and Figure 18. The non-acclimated seed showed no sign of surviving the initial depression in pH levels (Figure 17), producing only minute levels of methane throughout the experiment (Figure 22), thereby causing total VFA concentration to rise to as high as 20 g/L (Figure 18), a level too high for degradation produced alkalinity to have an effect on pH. This can be seen from the stagnant pH, which stabilizes at or near 5.5 (Figure 17). While Veeken and Hamelers (2000) provide evidence that degradation stability is dependent upon pH and appears to be not directly related to total VFA concentrations, at least up to 30 g/L, this system was in fact inhibited by the VFA, leading to system failure as represented by complete cessation of gas production by day 30, despite the presence of non-degraded organics.

In contrast, in the acclimated system, total VFA production stabilized achieving a steady state at about 12 g/L (Figure 18) despite the fact that total biogas continued to climb due to continued organic degradation (Figure 23). This is explained by the fact that for the acclimated system, significant methane production began as early as day five and continued to increase throughout the experiment, achieving 33% and 57% methane by days 15 and 30, respectively (Figure 23). Despite the low pH during Day 4 (4.9, as shown in Figure 17) which falls below the assumed threshold level (5.5) of methanogenic viability (Bitton, 1994), a certain degree of the acclimated seed was able to survive this pH pressure, and with time and a corresponding elevation in pH, was able to proliferate, thereby effectively converting future production of VFA to methane and holding overall VFA concentration steady (Figure 18). With VFA held constant through methane conversion, additional alkalinity resulting from continued degradation further allowed the pH to rise to near neutral (Figure 17), enhancing hydrolytic kinetics and providing a feedback loop for overall enhanced conversion production.

## Conclusion

As Hedge and Pullammanappalil (2007) concluded, it is possible to generate a stable LBR operating on a certain degree of highly biodegradable food/green waste using acclimated leachate effluent and/or acclimated seed. In this case 16.7% of the solid biomass was acclimated seed which is comparatively small in comparison to commercial LBR systems that use as much as 50% recycled biomass as seed (Lehtomaki and Bjornsson, 2006). However, despite achieving a stable operation, the LBR did experience an early rapid decline in pH to as low as 4.9, requiring 5-6 days of reaction to achieve VFA concentration equilibrium via established methanogenic activity and significant methane production. In addition, greater than 45 days of LBR operation were required to fully digest the biodegradable portion of the solid waste, all conditions that could perhaps be improved upon with further engineering design.

# **Section 7: Batch digestion using both internal leaching bed reactors and external upflow anaerobic sludge blanket feed**

## **Background**

The subsequent set of experiments aimed to reduce the pH depression seen in the earlier internal recycle batch runs, enhance the degradation kinetics and reduce overall reaction time required within a LBR. As discussed in Section 3, to accomplish this, external feed of UASB effluent to the LBR was carried out for a proscribed period of time prior to switching to only internal recycle. To that end, the experiments described in this section were conducted to ascertain the effect this external feed would have on overall operation and to determine optimal effects of external feed rate and alkalinity on initial pH depression. All of these experiments used non-acclimated seed and liquid.

While limited project time precluded the completion of a full statistical design to determine optimal conditions for feed rate and alkalinity, several batch runs were completed at three feed rates and two alkalinities. The external feed experiments were conducted well past the expected five-day window for external operation so as to better determine both the short and long-term effects of external feed on degradation and VFA accumulation. The experimental runs completed were as follows:

### **Effect of external feed rate with constant effluent alkalinity (4 g/L CaCO<sub>3</sub>)**

- *External Feed 4x of LBR Volume*—food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in UASB treated effluent (5 L) operated with a continuous 4x external feed rate of UASB treated effluent (20 L/day) while also maintaining an internal recycle rate of 5 L/day for the entire system.
- *External Feed 2x of LBR Volume*— food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in UASB treated effluent (5 L) operated with a continuous 2x external feed rate of UASB treated effluent (10 L/day) while also maintaining an internal recycle rate of 5 L/day for the entire system.
- *External Feed of 1x of LBR Volume*—food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in UASB treated effluent (5 L) operated with a continuous 1x external feed rate of UASB treated effluent (5 L/day) while also maintaining an internal recycle rate of 5 L/day for the entire system.

### **Effect of alkalinity with constant external feed rate (2x of LBR volume)**

- *UASB Effluent with Alkalinity of 2 g/L CaCO<sub>3</sub>*—food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in UASB treated effluent (5 L) operated with a continuous 2x external feed rate of UASB treated effluent (10 L/day) while also maintaining an internal recycle rate of 5 L/day for the entire system.
- *UASB Effluent with Alkalinity of 4 g/L CaCO<sub>3</sub>* — food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in UASB treated effluent (5

L) operated with a continuous 2x external feed rate of UASB treated effluent (10 L/day) while also maintaining an internal recycle rate of 5 L/day for the entire system.

## Results and discussion

As discussed earlier, the batch run conducted using non-acclimated seed or effluent (UASB granules and effluent (2 g/L alkalinity)) resulted in a precipitous initial pH depression to as low as 3.9. Figure 24 shows that various rates of external feed resulted in a smaller initial pH depression though higher external feed rates did not always result in lower initial pH drops. The pH lows were 4.4, 5.6, and 5.3 respectively for 1x, 2x and 4x external feed rates (Figure 24).

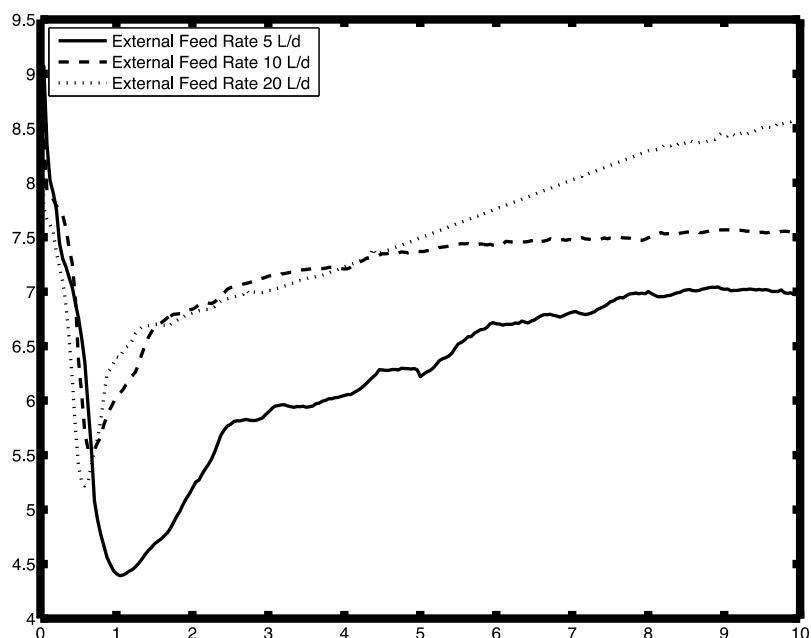


Figure 24. pH against reaction time for 1x, 2x, and 4x feed rates each at 4 g/L alkalinity.

Notably, two factors were altered from the baseline batch experiment: insertion of external feed liquid and insertion of liquid with twice the alkalinity as that which was originally loaded into the reactor. Therefore the noted increase in minimum pH achieved compared to the baseline experiment could be a result of higher initial alkalinity, a higher flow through rate allowing for more produced VFA to be removed from the reactor, and/or continuous insertion of suspended bacterial biomass present with the external liquid being fed. Figure 24 results, with each pH depression being higher than the earlier batch run, do indicate that flow rate can have an impact on the reactor pH, presumably through higher liquid flow rates which allow for a more rapid removal of potentially inhibitory VFA out of the reactor. However, as the optimal feed rate in terms of pH depression results was 2x, it can be seen that use of too high of feed rates can in fact minimize the gains, presumably through a washout of alkalinity and/or bacterial biomass.

Figure 25 results summarize the VFA within the reactor liquid and again show that higher feed rates do not necessarily achieve improved results, indicating maximum external feed or dilution rates to be utilized. The total VFA concentrations nearly doubled at the 2x feed rate compared to

the 1x feed rate, clearly showing that despite a greater liquid dilution, enhanced biological degradation was taking place.

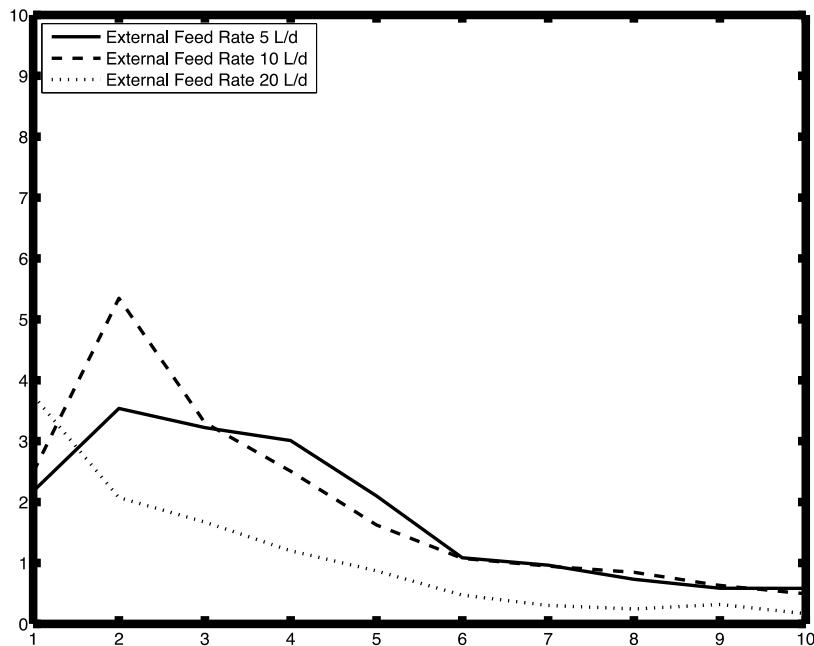


Figure 25. VFA against reaction time for 1x, 2x, and 4x feed rates each at 4 g/L alkalinity.

However, when the feed rate was further doubled to 4x, the total VFA concentration significantly decreased to levels that were below even that of the 1x run. While the greater liquid transfer rate of 4x might have accentuated the biological degradation, the increased liquid dilution overwhelmed any gains, again pointing to external feed rates that need to be optimized for both enhanced degradation kinetics and minimized for dilution effect.

Earlier batch trials for non-acclimated and acclimated conditions yielded biogas productions by day five of 25 and 90 L/kg VS loaded, respectively. During the external feed trials, biogas production was significantly enhanced over non-acclimated batch internal recycle only but only equaled the production rate of acclimated batch with the 2x feed rate (Figure 26).

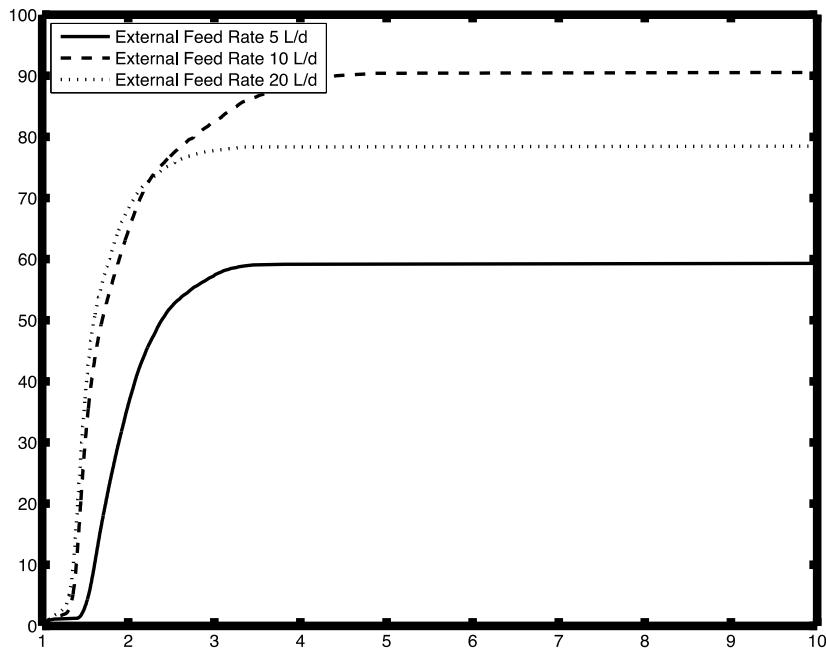


Figure 26. Biogas accumulation against reaction time 1x, 2x and 4x feed rate and 4 g/L alkalinity.

Put another way, batch operation with an external feed rate of 2x equaled the performance of the internal recycle only batch run with acclimated seed and liquid despite introducing liquid with a lower alkalinity (4 g/L to 5.6 g/L) and theoretically washing some established bacterial biomass out of the reactor. To achieve similar gas production, the external feed must have allowed for improved pH (5.6 to 4.9) through a combination of enhanced mass transfer and degradation kinetics, flow release of accumulating VFA, and return of suspended bacterial biomass.

Interestingly, not only is the total biogas production similar, but the component structure is approximately the same, with both runs producing about 10% methane by day five and 24% by day ten (compare total biogas production and methane in Figure 27 and Figure 23). Obviously at this external feed rate, alterations to biomass quantity and structure (insertion of suspended UASB granules enriched in methanogens and/or potential washout of acclimated reactor seed) were either minimized or had minimal effect on overall performance.

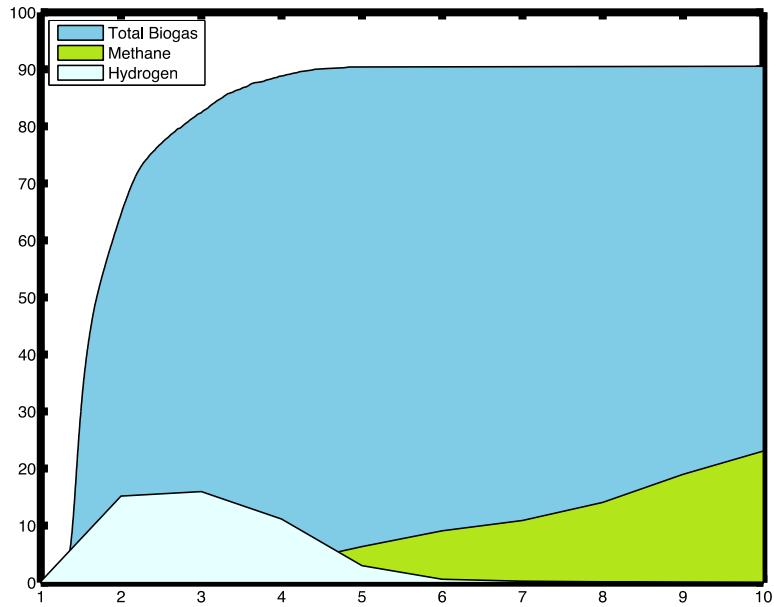


Figure 27. Biogas accumulation against reaction time with changes in biogas constituency for 2x feed rate and 4 g/L alkalinity.

One effect inclusion of an external feed rate at the beginning of the batch run has on reactor operation is that total VFA concentrations stay considerably lower than those observed during internal recycle with no external feed. Previous acclimated internal recycle only batch runs achieved roughly 14 g/L total VFA by day five, while the 2x feed rate of this set of runs yielded roughly 4 g/L total VFA through day five. This overall lower VFA concentration could have an impact on downstream operation of a UASB for treatment of the overflow leachate both in regard to handling the total concentration of acidic and suspended solids as well as the form of the acid components. Operation at dilute VFA concentrations induces acetate percentages of approximately 50% with that fraction gradually increasing as dilution is increased (Figure 28). Comparison to the earlier acclimated batch run with high VFA concentrations shows acetate fractions of 30-40% (Figure 20), meaning a greater fraction of easily digested acetate entering the UASB.

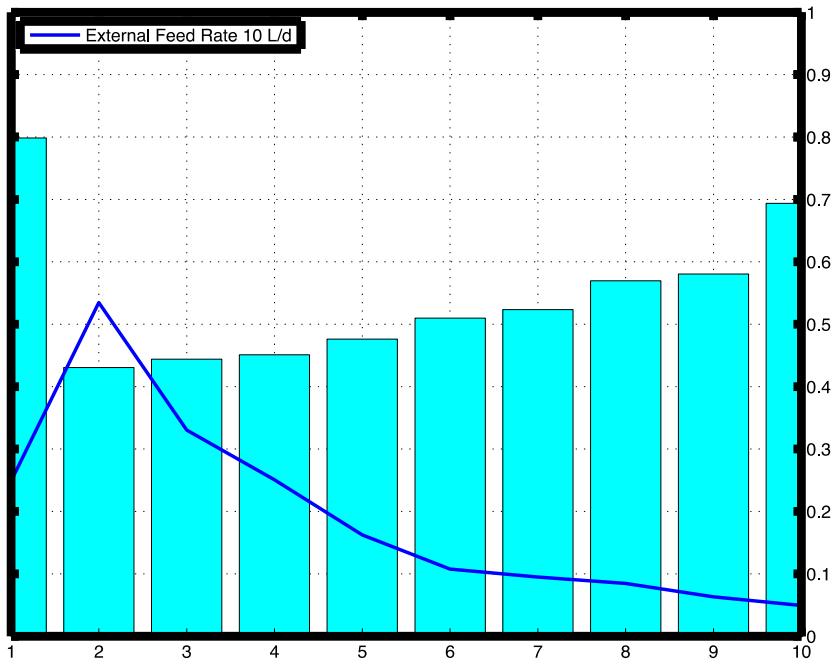


Figure 28. VFA against reaction time with change in percent acetate for 2x feed rate at 4 g/L alkalinity.

A series of additional external feed runs were completed to better ascertain the role of alkalinity in the process and to better determine the potential process enhancements that could be achieved through initial external feed and dilution. While the earlier discussion was related to various dilution rates using 4 g/L alkalinity, Figure 29 through Figure 32 summarize the effect changing alkalinity (4 v 8 g/L) on pH and total VFA when operating at respective 1x and 4x dilutions. Conclusions that can be drawn from these graphs are:

- Figure 29 and Figure 30 show that alkalinity play a major role in stabilization of early stage pH, significantly raising the minimum pH achieved. This is especially true at low external feed rates (compare the difference in Figure 29 to the difference in Figure 30). In turn, this can be expected to induce enhanced degradation kinetics due to operation at more preferred pH ranges. At high external feed rates, the gains in pH (Figure 30), even with high alkalinity feed, can be overwhelmed by effects of enhanced mass transfer, reaction and degradation kinetics, which produce significantly higher levels of total VFA that in turn impact pH (Figure 32).
- For the LBR system evaluated in this project, use of external feed at moderate rates, do lead to notable improvements in degradation kinetics, presumably through improvements in mass transfer.
- There are relatively complicated relationships between alkalinity, feed/dilution rate, mass transfer/reaction kinetics, and bacterial biomass. These relationships make attainment of preferred 7.0 pH ranges while also maintaining other desired outcomes problematic. Generally, initial external feed to the LBR can help attain the desired pH, but to only a degree.

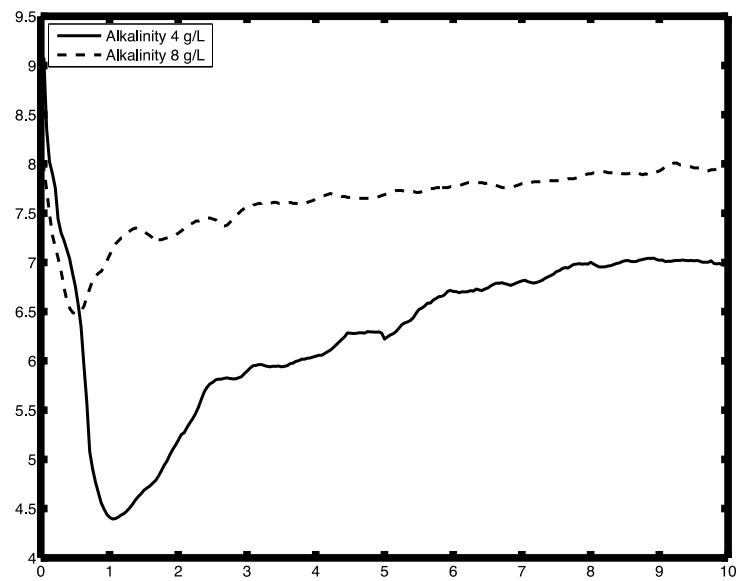


Figure 29. pH against reaction time for 4 and 8 g/L alkalinity, each at 1x feed rate.

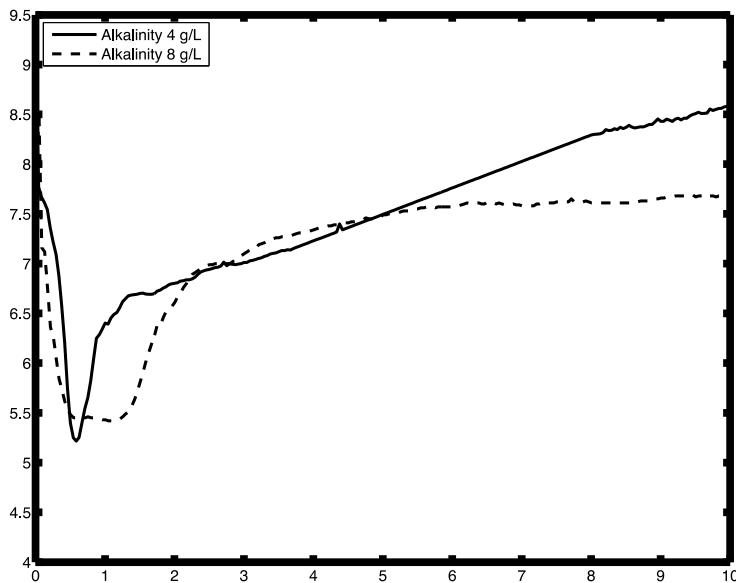


Figure 30. pH against reaction time for 4 and 8 g/L alkalinity, each at 4x feed rate.

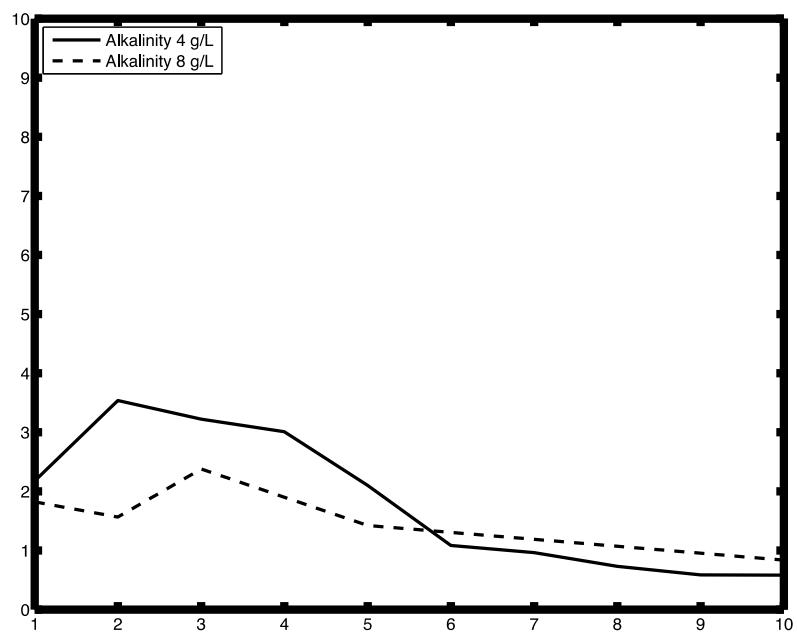


Figure 31. VFA against reaction time for 4 and 8 g/L alkalinity, each at 1x feed rate.

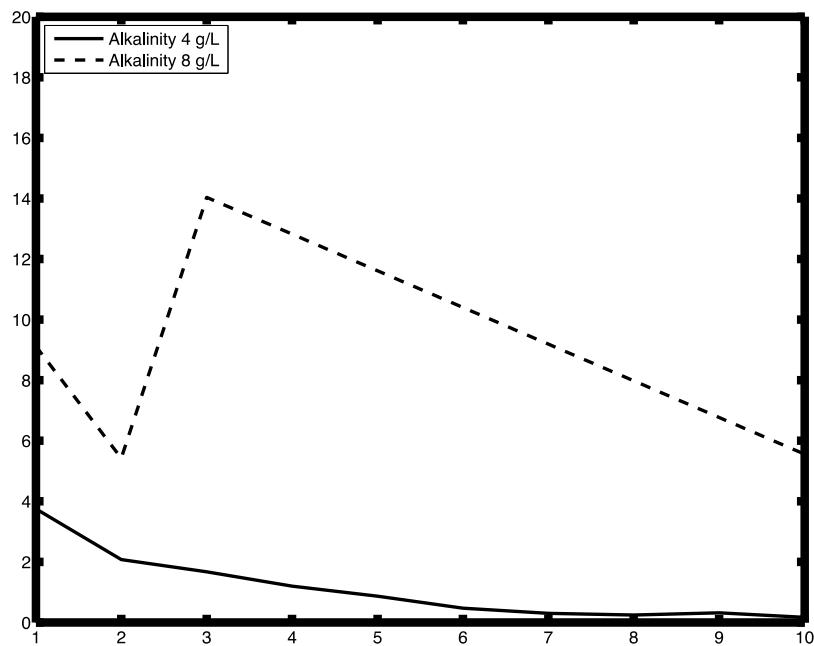


Figure 32. VFA against reaction time for 4 and 8 g/L alkalinity, each at 4x feed rate.

## **Conclusion**

Use of high alkalinity feed at high but moderated rates of feed achieved the best results. This was achieved through a combination of higher pH (enhanced hydrolysis and acidification), improved mass transfer (enhanced hydrolysis and acidification), and dilution and removal of VFA metabolites (enhanced pH and improved UASB operation). These results indicate that concerns regarding early stage pH depression experienced with internal recycle only batch runs can be in part addressed through the use of external feed input and recycle. However, an eye must be maintained on the inter-related aspects of alkalinity, feed/dilution rate, mass transfer/reaction kinetics, and bacterial biomass. While no direct quantitative evidence confirms this, it appears that insertion of bacterial biomass to the reactor through continual external feed had little to no appreciable impact on performance. This presumably occurs because of a comparatively low mass balance of entering biomass as compared to the existing biomass within the reactor.

## **Section 8: Batch digestion using phased internal leaching bed reactors and external upflow anaerobic sludge blanket recycle**

### **Background**

With initial batch trials completed for both internal recycle alone and various forms of external feed, a series of new trials was completed for evaluation of the full proposed system—namely a batch LBR operation using an initial period of external feed of digested effluent followed by a concluding period of internal recycle alone batch digestion.

Again, while the project time frame precluded the inclusion of a full statistical design to determine optimal conditions for the complete system, two experimental runs were completed evaluating the effect of seed type using previously tested and preferred levels of external feed rate (2x, 10 L/day) and alkalinity (4 g/L). Trials were as follows:

- *Full System using Non-Acclimated Seed*—food/green waste substrate (2.5 kg) seeded and mixed with UASB granules (0.5 kg) and then fully immersed in acclimated, digested effluent from a previous LBR run (5 L) operated with a continuous 2x external feed rate of UASB treated effluent (10 L/day and 4 g/L alkalinity) for the first five days while also maintaining an internal recycle rate of 5 L/day for the entire system. After day five, external feed suspended and system operated with only internal recycle (5 L/day) until suspension of the trial.
- *Full System using Acclimated Seed*—food/green waste substrate (2.5 kg) seeded and mixed with previously digested biomass seed (0.5 kg; 16.7% total solids mass) and then fully immersed in acclimated, digested effluent from a previous LBR run (5 L) operated with a continuous 2x external feed rate of UASB treated effluent (10 L/day and 4 g/L alkalinity) for the first five days while also maintaining an internal recycle rate of 5 L/day for the entire system. After day five, external feed suspended and system operated with only internal recycle (5 L/day) until suspension of the trial.

### **Results and discussion**

While the goal was to have initial and sustained pH during batch digestion in the neutral 7.0 range to achieve known enhancements in hydrolysis and acidification rates, operation of the full system under the tested conditions could not avoid an initial pH depression, which in the best case was 6.5 (Figure 33). While somewhat below the target, the 6.5 level was considerably better than that achieved by internal recycle alone (4.9). Also, Figure 33 confirms the contribution acclimated seed (biomass versus UASB granules) can make to maintenance of a preferred operating pH (6.5 for acclimated seed vs. 5.2 for non-acclimated). Lastly, this initial depression readily rose to a sustained neutral pH or higher (Figure 33), indicating fairly quick achievement of preferred operating pH.

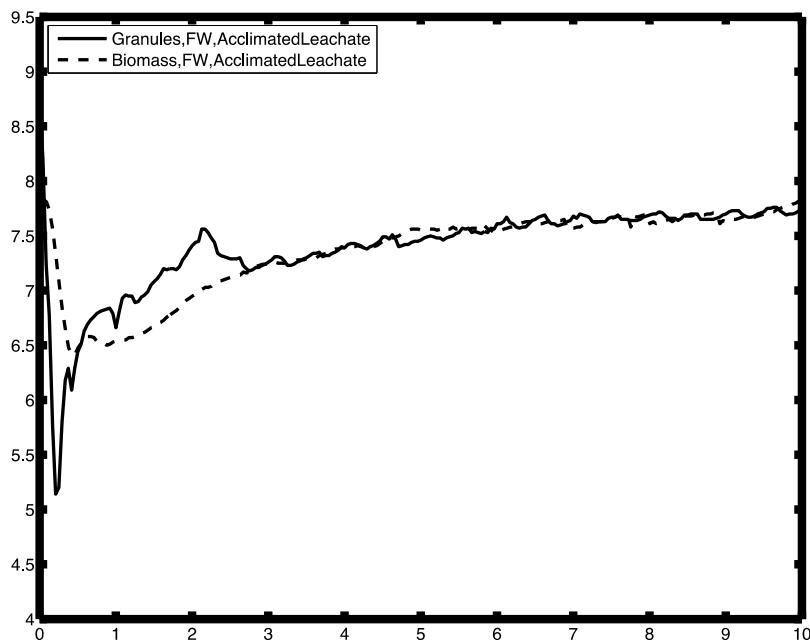


Figure 33. pH against reaction time for full batch system using either acclimated or non-acclimated seed.

Figure 34 summarizes the effect a full system operated with acclimated seed can have on degradation kinetics and performance, yielding total VFA concentrations three times as high as that of non-acclimated UASB granules during the initial five days of external feed. Importantly in comparison to the internal recycle only batch run with acclimated seed, total VFA concentrations spiked to a maximum of only 6 g/L (as compared to 16 g/L for the internal recycle only) and became sustained at or near 1 g/L for the course of the continued digestion (as compared to 14 g/L), thereby strongly enhancing reactor stability. The VFA profile was also enhanced, with a greater proportion of VFAs comprising easily digested acetate (Figure 35 as compared to Figure 20; 50-60% acetate compared to 30-40% acetate for much of digestion), and VFA loading to the downstream UASB.

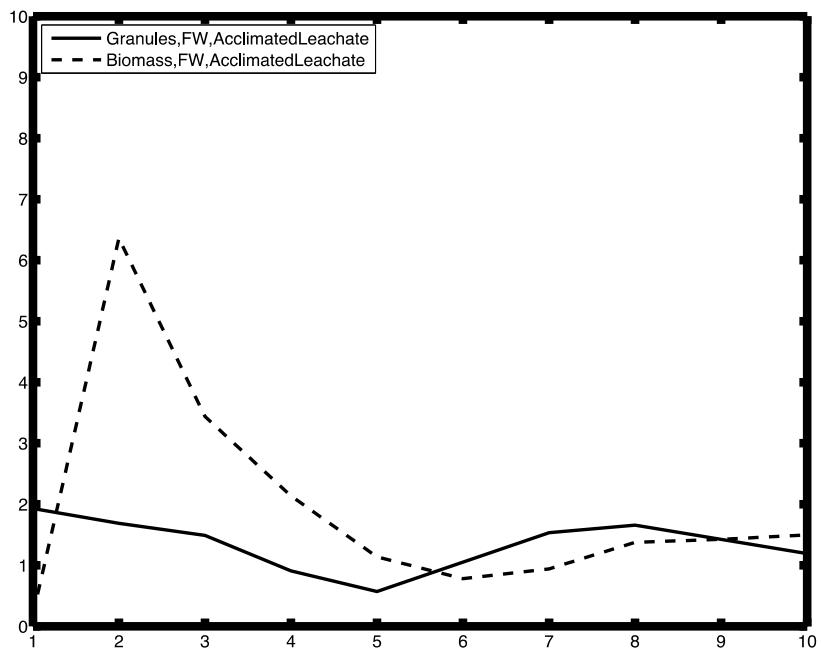


Figure 34. VFA against reaction time for full batch system using either acclimate or non-acclimated seed.

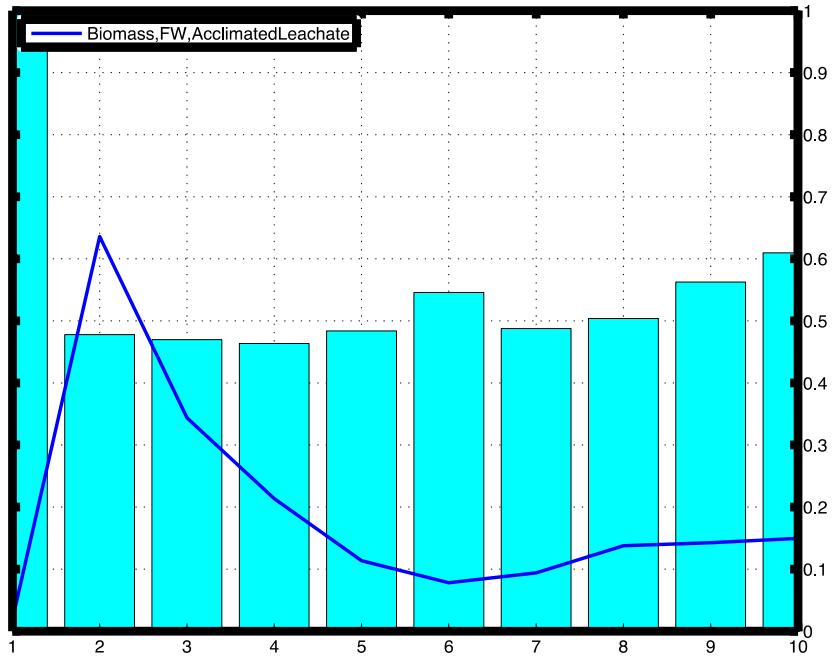


Figure 35. VFA against reaction time and change in percent acetate for full batch system using acclimated seed (16.7% biomass).

From a mass perspective, the internal recycle only batch run obtained VFA concentration of 16 g/L within a total volume of 5 L, producing 80 grams VFA by day two. This compares to the full system that spiked at 6 g/L VFA at the end of day 2 but contained 25 L of liquid, yielding a mass of 150 grams of VFA produced by the system or a 47% increase in degradation product. Clearly,

the full system and its use of external feed significantly enhanced mass transfer and overall kinetics.

While the impact on biogas production was less notable, operation of a full system using acclimated biomass seed was roughly 10% greater in overall production by day five or the stopping point of external feed as compared to use of non-acclimated UASB granules (Figure 36). When comparing the full system to the internal recycle only system, overall biogas production by day five were almost identical, each producing at day five 90 L biogas/kg VS loaded (compare Figure 36 to Figure 21). However, it is important to note that during these five days, the full system produced 50 L of excess effluent leaving the LBR for presumed downstream biogas treatment in a UASB, representing a significant portion of biogas that would allow the full system to significantly increase its biogas productivity as compared to the internal recycle batch control.

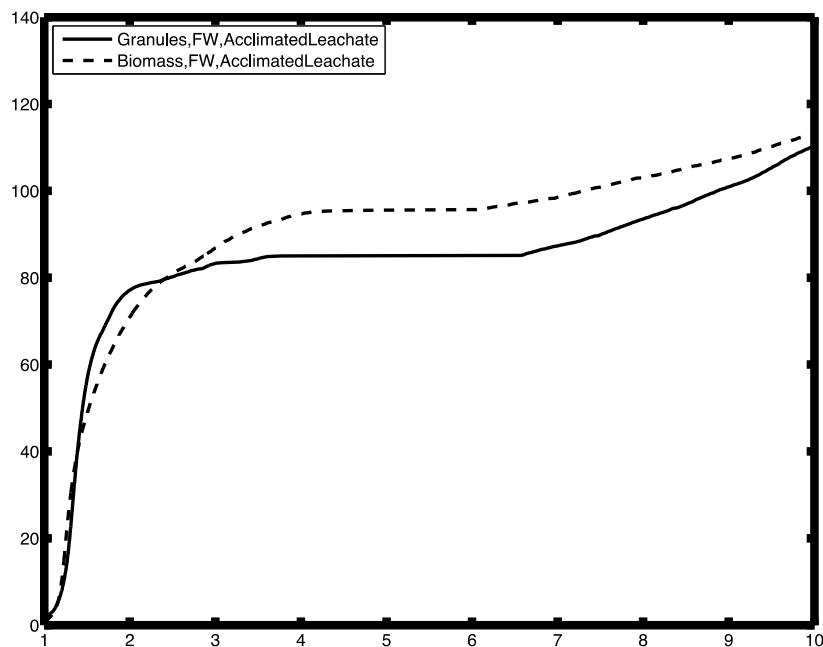


Figure 36. Biogas against reaction time for full batch system using either acclimated or non-acclimated seed.

Comparisons of Figure 37 and Figure 38 show that while little difference exists between the methane concentration profile for non-acclimated and acclimated full system batches, the acclimated seed produces significantly less hydrogen during the first days than the non-acclimated system.

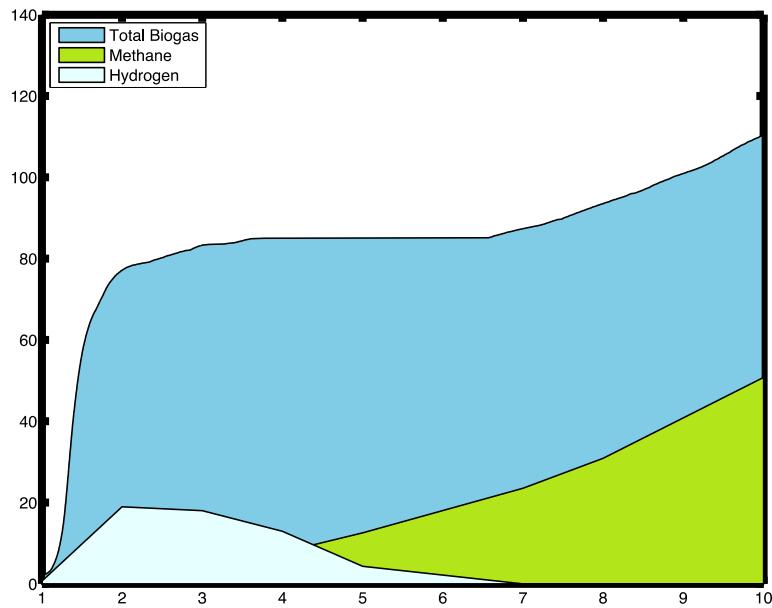


Figure 37. Biogas against reaction time and changes in components for full batch system using non-acclimated seed (UASB granules).

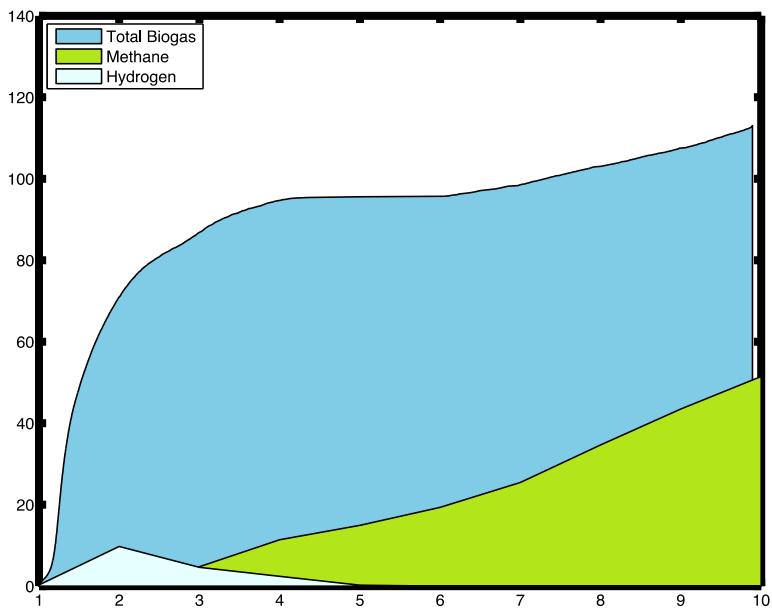


Figure 38. Biogas against reaction time and changes in components for full batch system using acclimated seed (16.7% biomass).

Hydrogen production is important as methane is currently the preferred end product, given that collection and purification of hydrogen (as is envisioned in some digestion systems producing large amounts of hydrogen gas) is still quite non-economical. The limited amounts of hydrogen

produced during the first stages of the external feed to the LBR could presumably be easily shunted to the envisioned down-stream UASB reactor for conversion to methane for even greater methane productivity as opposed to targeting the hydrogen as an end product.

Most importantly, Figure 39, comparing the methane composition of biogas in the full and internal-recycle only systems using acclimated seed, clearly shows the advantage of the full system. The percentage of methane in the biogas is much higher after the initial few days for the full system than the internal recycle alone system. For the full system, regardless of seed utilized, the methane percentage for days five to ten and then twenty rises from 15.6 to 45.5 to 68.7% while internal recycle alone starts at roughly 12.8 and rises to 25.5 and 46.0%. While not directly measured, these higher percentages of methane would presumably favor the development of a more robust population of methanogens during the time of system establishment, enhancing system productivity. Clearly, use of the full system with its initial external feed for pH control and enhanced kinetics has an impact on methane production.

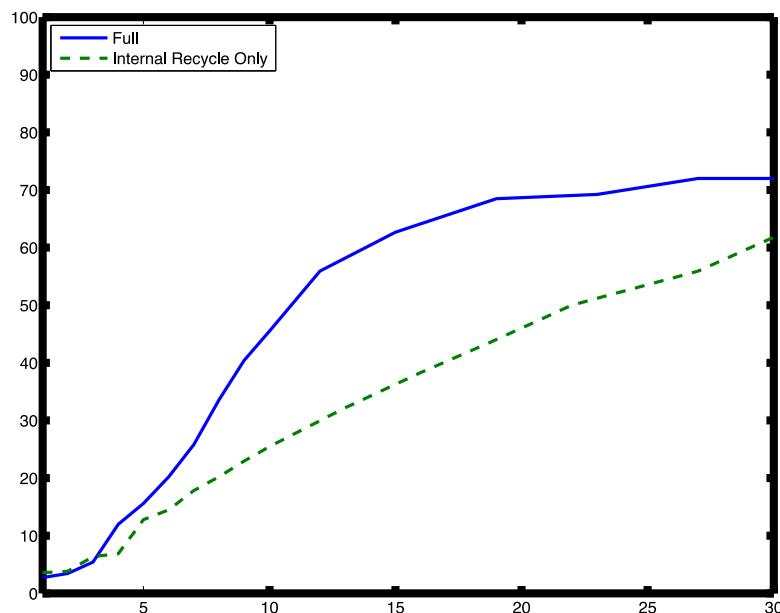


Figure 39. Comparison of methane percentages against time for full and internal recycle only systems.

## Conclusion

For both the full and internal recycle alone systems, acclimated seed notably benefitted reactor stability and performance. Of greater importance to this project, incorporation of a full system protocol merging initial external feed with internal recycle also led to improvements in reactor stability and performance. As compared to an internal recycle alone batch digestion in an LBR, the full system allowed for these changes and enhancements:

- Higher initial pH operation capable of achieving minimum levels that did not drop below 6.6, allowing for operating ranges more conducive to effective hydrolysis and acidification.

- Greater mass transfer and kinetics during early period of digestion, yielding 47% increase in VFA production by day two.
- VFA concentrations with lower maximum and sustained values that are more conducive to down-stream UASB digestion.
- Reduced levels of hydrogen gas produced during initial stages, allowing for more ready incorporation to UASB stream and conversion to methane.
- Significantly more rapid increase in methane percentage over time and therefore presumably increased methanogen activity—achieving 45.5% methane by day ten as compared 25.5% for internal recycle alone.

# Section 9: Full system performance

## Background

While Section 8 results suggest that the full system provides important process improvements compared to a baseline internal-only leaching system, this section focuses on the full system's potential capabilities in regard to mass destruction, biogas production and nutrient transfer. Note that this discussion is based on the non-optimized full system runs completed and discussed in Section 8. As already noted, the LBR system used in this project was a submerged up-flow system, which could differ from true non-submerged leaching bed designs in regard to its mass transfer and kinetic capabilities. Therefore, comparison of this project's full system capabilities to commercial systems should be made cautiously, especially in regard to comparisons of volumetric productivity. In some (but not all) respects, differences between the experimental system and a non-submerged leaching bed system would actually lead one to expect that the non-submerged system would have improved performance in regard to some performance measures, particularly in the required days for preferred solids destruction given the higher mass transfer capability of the non-submerged system.

## Results and discussion

Table 6 below summarizes the solids destruction and nutrient movement during operation of the full system as designed and tested. During 37 days operation of the LBR (bed of 2.5 kg food waste containing 0.5 kg seed and submerged with 5 L digested effluent operated under protocol of 5 days of external feed at a rate of 10 L/day and internal recycle rate of 5 L/day followed by only internal recycle at 5 L/day) 69 and 70% of TS and VS respectively were either biologically destroyed directly to methane or transferred out of the reactor into the external overflow destined for downstream treatment in a UASB-type reactor. Using earlier proximal analysis, the theoretical methane productivity of the food/green waste loaded to the LBR is known to be 560 L CH<sub>4</sub>/kg VS and since the LBR produced 128 L of CH<sub>4</sub>, then presumably, 228.6 g of VS were directly destroyed to methane in the LBR. This value represents 44% of the total VS lost from the LBR, thereby confirming that under this design, the LBR acts as much more than a leaching hydrolysis/acidification reactor and importantly equally so as a methane producing reactor.

Table 6. Mass balance of full system for LBR and external UASB components (37 days).

	TS (g)		VS (g)		TN (g)		TAN (g)		TP (g)	
	LBR	Ext	LBR	Ext	LBR	Ext	LBR	Ext	LBR	Ext
In	794	262	746	9.0	25.1	4.6	3.1	1.6	2.8	2.5
Out	247	703	227	449	13.9	15.8	<0.1	13.4	2.6	2.7
Reduce (% or x)	69	-2.7x	70	-49.9x	44.6	-3.4x	>97	8.4x	7.1	8.0

Table 6 also gives some indication of the degree to which nutrients are converted to inorganic or soluble form and transferred to the external liquid leachate. For total nitrogen (TN), total ammonia nitrogen (TAN) and total phosphorus (TP), 44.6, >97, and 7.1%, respectively were transferred out of the solid pile in the LBR to the leachate, allowing for more ready and economical nutrient recovery from this liquid phase. Importantly, during the biological processing, a considerable fraction of the nitrogen was mineralized and converted from organic to inorganic form, with ammonia mass levels in the leachate increasing from 1.6 to 13.4 grams (a 8.4 times increase).

The above nutrient transfers and transformations have profound implications for downstream processing within an assumed organics recycling yard. As Section 1 indicated, many compost facilities that are beginning to treat a higher percentage of putrescent food/green waste are experiencing odor concerns at their facilities, including the release of ammonia during the active biological aeration process. AD pretreatment will significantly reduce the levels of organic volatiles and total nitrogen (44%). With these lowered levels of organics and nitrogen, processing of the solid residuals in compost piles could conceivably occur with less concern about release of ammonia or other volatile odorous compounds.

Nitrogen could also presumably be feasibly recovered from the liquid with existing technologies, given that it has been converted to ammonia and is in an easily flowing, low-solids, concentrated state. Table 7 tracks the flows and form of nitrogen throughout the process. With similar transformation, a facility that processed 100,000 MT/year would potentially produce 400 MT of nitrogen with an assumed 70% ammonia recovery efficiency. Importantly, ammonia concentrations within the recycled digested leachate for re-use in new LBR operations would also be maintained at low concentrations with this method, removing concerns that ammonia could build to inhibitory levels through sustained operation, as is the case with repeat batch operations not involving nutrient recovery. Optimization of the full process should aim to produce the highest possible ammonia concentrations in the leachate, which could result in improvements over these initial performance calculations. Presently, with five days of external feed at a 2x flow rate, the leachate has concentrations of ammonia in and around 0.3 g/L.

Table 7. Nutrient flows.

	Conc. (mg N/L)	TAN (%)	g/wet kg
Residual Solids (before composting)	880,000	Negligible	5.6
LBR Leachate	280	85	6.3
Digested Leachate	280	91	6.3
Return (70% rec. TAN)	102	75	2.3
Recovered Nutrients from Leachate	178	100	4.0

Table 8 is a summary of the gas production efficiencies of the full system as tested. Nearly 46% of the methane production occurs within the LBR, which primarily operates as a methane reactor as opposed to merely a hydrolyzing/acidification reactor, through maintenance of a 6-7 pH range, and rapid production of methane (nearly 20% methane composition by day five and cessation of external feed).

Table 8. Gas production performance indicators for full system by component and total (37 days).

	Biogas		Methane + H <sub>2</sub>	
	LBR	UASB	LBR	UASB
Out (L)	258	235 <sup>a</sup>	128 + 11 <sup>b</sup>	153
SMP (L/kg VS in)	359	524	178	341
SMP Total (L kg VS in)		685		390
TMP (L/kg VS) <sup>c</sup>		860		560
Biodegradability (SMP/TMP)		79.7%		69.6%
VMP Total (m <sup>3</sup> /m <sup>3</sup> day)		1.8		1.0
OLR (kg VS/m <sup>3</sup> LBR day)			7.8	
Total Reduction (% VS) <sup>d</sup>			66.8	

<sup>a</sup> assume 65% methane content from UASB reactor

<sup>b</sup> assumes that H<sub>2</sub> produced in LBR sent to UASB for conversion to CH<sub>4</sub> at 4:1 volumetric basis

<sup>c</sup> based on Bushwell equation from lipid, protein, carbohydrate and VFA components

<sup>d</sup> assume a VFA TMP of 400.7 L CH<sub>4</sub>/kg VS at 20C; 60/40% mix acetate/butyrate; system not mass balanced

Hydrogen production only occurs during the first five days, achieving a zero percentage by day five. With such a short duration of hydrogen production, it is envisioned that the biogas production from the LBR headspace be fed through to the accompanying UASB for conversion of the hydrogen to additional methane for enhanced productivity. Thus, in the test bed, 11 L of H<sub>2</sub> are assumed to produce at a 4:1 volumetric conversion rate, an additional 2.75 L of CH<sub>4</sub> to the system. On a percentage basis, conversion of this hydrogen fraction amounts to 1% of total production of methane, not a significant source but nonetheless effectively utilized.

Earlier discussion of theoretical performance assumptions given in Section 4 can now be compared to actual performance operation. In review, other literature indicated that a high solids digester could operate at an organic loading rate (OLR) of near 10 kg VS/m<sup>3</sup> day and yield a specific methane productivity of 340 L CH<sub>4</sub>/kg VS loaded while reducing total VS by 60% at a volumetric productivity of between 1-4 m<sup>3</sup>/m<sup>3</sup> day. Table 8 shows that final performance indicators were an OLR of 7.8 kg VS/m<sup>3</sup> day, yielding a specific methane productivity of 390 L CH<sub>4</sub>/kg VS loaded while reducing total VS by 67% at a volumetric productivity of 1.8 m<sup>3</sup>/m<sup>3</sup> day. These performance indicators place the system right in the middle of expected performance as indicated by Hartmann and Ahring (2006) and well within desired targets. As noted earlier, use of a submerged up-flow system made necessary for testing at this scale, could have perhaps reduced mass transfer efficiency and resulted in abnormally long reaction times which in turn affected overall reaction time required and the longer-than expected timeframe for cessation of biogas production. These longer reaction times depress specific methane and volumetric biogas productivities; thus changes to this flow regime could conceivably raise the performance standards to the higher end of the expected range.

## Section 10: Scale-up and conclusions

### Scale-up

Figure 40 is a summary schematic for a scaled version of the full system being proposed and tested in this study. The scale is 100,000 MT/year and is envisioned to be a part of an organics-recycling center with AD being a pretreatment option for the food scraps and some of the green waste entering the yard.

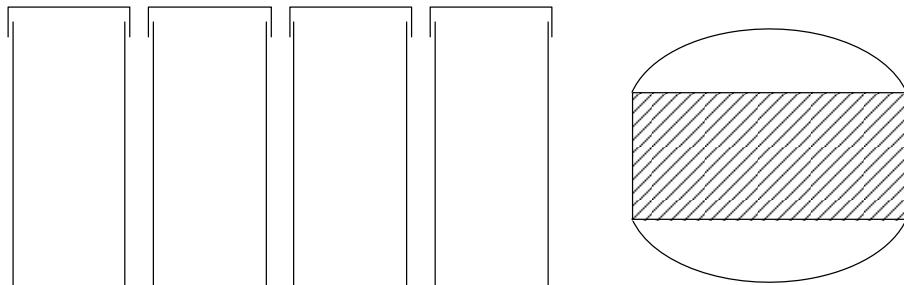


Figure 40. Scale up of single phase, dual reactor AD process for treatment of food/green waste.

The legend for the schematic is as follows, referenced to the numbers in Figure 40:

1. A 100,000 MT/year facility will require 9 trucks/day with an assumed truck volume of 40 cubic yards or 30.6 MT.
2. This amounts to 275 MT/day of 70/30% food scrap/green waste coming to the facility to be

treated in the AD system. Of course additional organics will enter the facility gate for treatment directly in the compost yard. Various types of AD pretreatment might be completed prior to entry into the digester, such as maceration, mixing, etc. although the batch system being designed is tailored for the minimization of needed pretreatment.

3. The system is designed to have 20 LBR bays, each capable of being loaded with and treating 550 MT food/green waste and handling 1,200 MT of liquid during the leaching process. The reactor volume required for each bay to handle this inflow, with an assumed porosity of waste material of 50%, is  $1,425 \text{ m}^3$  with a dimension of 20' x 20' x 125'. The design SRT is 40 days based on earlier project findings, however use of varying leaching approach, i.e. non-submerged, down-flow, could potentially improve upon mass transfer and decrease the required SRT. For purposes of this design the operating temperature is 37.5 °C.
4. During the first 5 days of operation, UASB treated effluent with high alkalinity and low levels of organics and nitrogen will be fed to the LBR as external feed to modulate the reactor pH, mass transfer, hydrolysis/acidification kinetics, and product-inhibition risk. The flow rate is at 2x the holding volume of the LBR (2,400 MT liquid/day). There is also internal recycle pumping at 1x the holding volume (1,200 MT liquid/day) to maintain leaching behavior and mass transfer. Excess liquid from the external feed will overflow to a buffer tank prior to treatment in the UASB.
5. A UASB is envisioned as the high biomass, high efficiency liquid digester, although various other biomass retaining, high rate digesters could be used. In order to treat the external flow rates described above from a single LBR as well as a portion from two other LBR that are on a snap-shot 5 day sequential processing timeframe, the UASB will need to be able to process  $3,600 \text{ m}^3$  of leachate/day, requiring a working UASB volume of  $5,130 \text{ m}^3$  based on typical design parameters for commercial UASB and an assumed HRT of 0.5 days.
6. After treatment with the UASB, digested effluent is passed through a nutrient recovery unit with an assumed capability of removing 70% of ammonia N in the effluent. The assumed process is an ammonia stripping process, but other nutrient recovery processes could be used as well, such as struvite, etc. No dimensions, costs, etc. are predicted for this assumed black box operation.
7. It is this treated liquid, digested and with nutrients recovered, which is used as external feed to the LBR for the first five days as earlier mentioned. After 5 days, the external feed is shut off with that particular LBR operating the rest of its SRT in internal recycle mode only. There is a mechanism for the headspace gas from the LBR to be sent to the UASB via gas recirculation during the first 5 days so that H<sub>2</sub> gas produced during the initial stages can be further processed to methane by methanogens active in the UASB.
8. After the 5 days, the LBR will be in internal recycle mode. At the completion of the planned 40 day SRT, the bay will be opened, liquid will be stored and utilized as initial liquid load to a new LBR as well as 16.7% by mass of the residual biomass for use as bacterial seed. The internal recycle will be at 1x the original load liquid volume.
9. Residual solids can then be either directly used as soil amendment (as near complete

degradation has taken place) or additional curing can be accomplished through incorporation of the residual solids within compost piles. The residual solids can also represent material to mediate and/or seed compost piles for optimal C/N ratio, etc.

10. Biogas and therefore methane from the LBR and UASB headspace will be sent to an engine/generator set for production of combined heat and power (CHP), although the gas could instead be scrubbed for processing into compressed transportation fuel and/or pipeline gas. Assuming typical engine/generator efficiencies of this size ( $9.4 \text{ ft}^3 \text{ CH}_4/\text{kWh}$ ), the facility should be able to gross produce nearly 4.8 MW of electricity as well as plenty of recovered heat for maintenance of digester heating needs while still having excess process heat for valued use in or near the organics processing facility.

## Project conclusions

Below is a summary of project research conclusions to be drawn from this study:

- A batch LBR can be operated at sustained high pH levels, with minimal early depression, if a protocol of external leachate feed is introduced to the LBR during the initial days. This enables production of notable amounts of methane in both reactors of this single-phase, dual reactor, batch LBR system.
- Experimental evidence shows that this early stage external feed protocol positively impacted pH maintenance/elevation, mass transfer and biodegradation kinetics. Experiments did not quantitatively ascertain extent to which these improvements were related to either improved mass transfer, preferred biological operating pH, or enhanced bacterial biomass (through incorporation of inflow bacteria). Analysis of the results alongside a literature review suggests that the primary mechanism was maintenance of preferred pH range for more optimal bacterial kinetics, followed closely by improved mass transfer with the additional material flow. Insertion of bacterial seed appeared to have the least impact, presumably because the inflow bacterial mass was small in comparison to existing bacterial mass developing within the LBR.
- Although optimal values were not determined, data shows that external feed protocols with high alkalinity and moderate flow rate were preferred. In this system, available alkalinities from the LBR/UASB system appear to equilibrate at near  $5 \text{ g/L CaCO}_3$ , making this the practical maximum to be fed to the system given this particular feedstock composition. Low feed rates, such as 1x the reactor liquid volume, did not allow for proper improvements in mass transfer from both a biological activity and product-inhibition removal perspective while too high of feed rates (i.e. 4x or above) led to too much dilution, and washout of developing alkalinity and/or bacteria.
- While other commercial versions of batch LBR digestion operate using high levels of biomass recycle, nearing or exceeding 50% of previously treated residual solids loaded as seed to new batches, this system operated effectively at 16.7% inclusion. Use of a greater seed ratio and/or changes to leaching operation (i.e. non-submerged, down-flow) could potentially improve upon degradation kinetics and allow for shorter SRT within the LBR, however, conclusive evidence for this was beyond the scope of this study.

- Incorporation of the external feed allowed for additional system improvements, including production of leachate VFA concentrations, VFA component profiles, and pH more preferred for treatment within a liquid, high rate reactor. The system's feed was at VFA concentrations near 3 g/L, comprised of a greater percentage of acetate and all at a pH near 7 (compared to leachate VFA concentrations bordering on 13 g/L, comprised of large percentages of butyric acid and at depressed pH (~4-5.5) for a similar system lacking external feed).
- The combination of external feed followed by internal recycle also supplied other benefits. A short duration of external feed as opposed to continued external feed minimizes the liquid recycle, pumping and UASB volume needs. Although not optimized in this study, the length of external feed could be potentially reduced without fear of digestion upset, thereby further reducing the operating costs to the system. Use of a primarily internal recycle regime allows for development of a final liquid batch high in bacterial seed and alkalinity which is ideal for new LBR loading and operation, contributing to a reduction in required residual solids for seed.
- Operation of the LBR also allowed for more ready cessation of H<sub>2</sub> production and earlier production of CH<sub>4</sub> as compared to no external feed batch controls. By generating CH<sub>4</sub> faster and to higher level sooner (nearly twice the percentage of CH<sub>4</sub> for a given time period during initial periods of LBR operation), the external feed system created a feedback loop which further controlled pH and VFA concentrations. In the end the single phase, dual reactor approach produced 46% of its methane in the LBR.
- By minimizing the period in which H<sub>2</sub> was produced, a mechanism for injecting this headspace gas to the UASB could allow for ready conversion of H<sub>2</sub> to additional CH<sub>4</sub>, thereby improving overall process capabilities, although data showed that the impact was merely an overall 1% adjustment to total methane production.

While use of a single phase, dual reactor approach brings added complexity and therefore the potential for elevated capital and operating costs compared to simple batch, internal recycle-only AD system, the above study does show promise on several fronts. Productivity comparisons to commercial systems that are included in this study should be taken with some degree of caution, as throughout the study, protocols, particularly the mechanism for leaching and mass transfer, were not optimized. Primary focus of the study should rest on the proof of the concept regarding development of a batch LBR system that can function as a single-phase dual reactor system with discussed potential for process improvements.

## Part II: Moderate Solids System

This project, the second of two research projects on anaerobic digestion carried out by Washington State University (WSU) for the Washington Department of Ecology (Ecology) Waste 2 Resources (W2R) program, constructed and investigated a floor-scale anaerobic digestion system for moderate solids (~15%) wastes such as food wastes. Ecology estimated that in 2008/2009, Washington State, with a diversion rate of 55%, landfilled nearly 5 million tons of MSW (WA Ecology, 2010a; WA Ecology, 2010b), with 18% by weight being food scraps, highly appropriate for anaerobic digestion in a moderate-solids system.

The moderate solids system included a high solids digester and a second up-flow anaerobic sludge blanket (UASB) reactor, with relatively high liquid recycle rates from the second unit to the first unit. This allows effluent from the second reactor to seed the first reactor with methanogens, while retaining high levels of methanogens in the second stage digester. At the same time, the process maintains the pH at 6.0-7.5 in both reactors, making methane production possible during both phases. Floor-scale experimentation with this novel two-reactor technology was merited based on prior results from laboratory-scale experiments, simulation and an economic analysis (Zaher et al., 2009). Together, these previous results had demonstrated the potential for the technology to out-perform current commercial technologies in terms of biogas production, biological stability, and construction and economic costs (Zaher et al., 2009).

Through the current research project, three main contributions were made to improve our new HSAD technology's competitiveness and scalability. The UASB reactor improved biogas productivity. In addition, two design improvements have been demonstrated in this project, which previous research suggests could cut mixing costs by more than half while also simplifying the system, (Zaher et al., 2009). First, an effective mixing strategy has been developed that combines three existing technologies in an energy-efficient manner: intermittent mixing, a design that sprays the effluent from UASB seed reactor into the high solids digester at the top inlet, and natural mixing arising from the force of biogas bubbles rising to the top of the high solids digester. Second, natural forces of the ongoing anaerobic digestion has been used to separate liquids and solids, avoiding the need for a more costly separation technology that would otherwise be necessary for the moderate solids digestion.

A floor scale system that utilized natural mixing and incorporated other design improvements was operated stably over a six month period with food waste feedstock. The in-depth knowledge obtained from the floor-scale moderate dual reactor system provided information on preferred operational conditions that will contribute to future pilot-scale demonstration efforts and eventual commercial-scale application. The high solids digester maintained operation at a pH of 6.0-7.0 (optimal for volatile fatty acid production and methanogen growth) without the need for additional buffering. Solids retention times (SRTs) were 7 days under batch mode operation and 30 days under continuous operation at a solids loading rate of 5.0 – 15 kg VS/m<sup>3</sup> day compared with WWTP loading rate of 1.91 – 2.16 VS/m<sup>3</sup> day (Ghosh et al., 1995). Hydraulic retention times (HRTs) were less than 1 day to separate liquid and solid in the high solids digester.

The UASB seed reactor provided adequate effluent to the high solids digester for mixing, pH control, and seeding of anaerobic microorganisms. It achieved up-flow velocities in the range of 4 - 10 meters per hour without washout of methanogenic bacteria, providing enhanced methane productivity compared to similar systems with other digester types (Seghezzo et al., 1998). The HRTs for the UASB were in the range of 6 to 24 hours. The organic volumetric loading rate for volatile fatty acid (VFA) was up to 15000 - 24000 mg/L at 35 °C. The pH was about 7.4 - 7.6. The methane ( $\text{CH}_4$ ) content of the biogas was 63.7%.

A process model that reflects the improved floor-scale moderate solids system has been developed to assist in the process of scale-up and future commercialization. The model, based on anaerobic digestion model No. 1 (ADM1) (a model that has been frequently used in research and commercial anaerobic digestion systems), includes both reactor and kinetic model components that can predict performance under a variety of sizes of larger-scale systems. Initial parameters in the kinetics model were determined using the default values of ADM1 and the model was then verified and adjusted using data from the floor-scale experiments. The model was used to examine the impact of changing the recirculation rate on process stability and kinetics, since this information is difficult to obtain from experiments. Results showed that a recirculation rate of 0.19  $\text{m}^3/\text{day}$  was optimal, maintaining pH at the range of 7.0-8.0 in the high solids digester. Under these conditions, the pH in the UASB seed reactor was maintained at 7.5-8.0 and most of VFA was consumed in this reactor.

Lastly, effluent from the floor-scale system was successfully treated in a laboratory-scale integrated, multifunction ammonia removal and nutrient recovery system (IMARNRS), previously developed at WSU. Results showed 93% TAN removal achieved over one day, and phosphate removal of 45% achieved over six hours. Adding nutrient removal capabilities would improve stability by removing high amounts of N and P that could inhibit the system. It also has a potential to generate salable byproducts that could improve economics and environmental benefits. Using experimental data, it was estimated that integrating IMARNRS into the effluent recycle loop of a facility processing 100,000 MT of food waste per year could produce 376 MT of nitrogen and 40 MT of phosphorus per year.

The results presented here support ongoing work towards scale-up of the new HSAD design. It is recommended that the three major design improvements incorporated in the floor scale system (use of UASB reactor, intermittent mechanical mixing, and liquid-solid separation technology) be continued in future scale up work. The floor scale work also suggested several additional design improvements that should be implemented in future pilot-scale work. Pumping biogas from the high solids digester into the UASB during initial system set-up would allow micro-organisms in the UASB to convert the high volumes of  $\text{CO}_2$  and  $\text{H}_2$  that are initially generated from the food wastes into methane. The use of a vertical liquid outlet in the high solids digester would effectively separate liquid and solid and avoid clogging the liquid outlet. Redesign to elevate and sequence operations, replacing pumps with gravity transport where feasible, would further cut capital and operational expenses.

WSU is currently discussing collaboration with potential industry partners to implement the system at the pilot scale. Goals of the pilot scale design would be to refine the engineering design and generate additional economic analysis relevant for commercial scale planning. Using data generated during laboratory scale work, plus a scaling factor to account for anticipated

changes during scale-up, it was previously estimated that operational and capital costs may be nearly a third lower than existing commercial HSAD applications (Zaher et al., 2009). It is thus estimated that using the new WSU design a facility processing 17,640 tons food waste per year could save greater than \$7.5 million annually, compared to an existing AD technology (Zaher et al., 2009).

## Section 1: Introduction

Part I provides detailed information on the amount of waste generated in Washington State each year and background on interest in high solids anaerobic digestion (HSAD). In brief, a US EPA (2008) summary of waste generation showed that in 2007, 54% of the 230 million metric tons (MMT) of municipal solid waste (MSW) produced in the US was disposed of in landfills. Meanwhile, in 2008, Washington State disposed of nearly 5 million tons of MSW with a diversion rate of 55% (WA Ecology, 2010a; WA Ecology, 2010b). Landfilled material contained about 27% organics and 18% food scraps, much of which would be highly appropriate for anaerobic digestion in a moderate solids system (WA Ecology, 2010b).

Representing 21% of total MSW production, yard trimmings have had a long history of being recycled and treated through aerobic composting with over 64% of total national production being diverted for treatment and eventual production of soil amendment products (US EPA, 2008). In contrast, the recycling of food scraps is quite immature with estimates of only 2.6% of a 28.8 MMT annual production recycled and treated on a national level, primarily through composting, leaving the overwhelming percentage of its organic material decomposing in landfills (US EPA, 2008). These food wastes would be highly appropriate for anaerobic digestion in a moderate-solids system.

In this project we constructed and investigated a floor-scale dual reactor system for food waste with a moderate solids concentration range of 15-20%. This digester system has a unique two-stage configuration, utilizing an upward-flow anaerobic sludge blanket (UASB) reactor as the second stage digester. The system configuration recycles high levels of flows from the second unit to the first unit (4-10 meters per hour) compared to existing systems, with little washout of the artificial media in the UASB reactor. This allows effluent from the second reactor to seed the first reactor with methanogens, while retaining high levels of methanogens in the second stage digester. At the same time the pH can be maintained at 6.0-7.5 in both reactors, making methane production possible during both phases. Mixing at the floor scale was accomplished through buoyant forces caused by self-produced biogas.

Results from previous bench-scale research had indicated that this technology was promising enough to warrant floor-scale study. Laboratory testing and modeling demonstrated that the system's loading rate and biogas production rate were improved by about 50% compared to existing HSAD designs (Zaher et al., 2009), while achieving comparable chemical oxygen demand and total solids reduction. Based on the results of previous laboratory-scale study, the cost of treating organic waste with the dual reactor system was estimated to cut current commercial treatment cost/kWh, as calculated by capital and operational costs amortized over the predicted life of the facility, by 30% to \$1.08/kWh compared to \$1.55/kWh calculated for an existing technology (Zaher et al., 2009). These numbers compare a lab-scale experimental design to actual commercial performance, but incorporate a scaling factor to compensate for this.

## Section 2: Project objectives

The two major goals of this project were to: 1) scale-up earlier laboratory designs for a moderate solids system to floor-scale and test the system, and 2) test and calibrate mathematical models to produce a model for a moderate solids system that reflects data from the floor-scale system. These goals were set to provide information relevant to a larger scale pilot system in the near term and facilitate the future commercialization of the moderate solids waste dual reactor system in the longer term.

Specific objectives included the following:

*1. Design and build floor-scale moderate solids system and optimize using laboratory-scale experimentation*

Design and build a moderate solids floor scale digester with design improvements based on the lessons learned in previous bench-scale research. Additional laboratory-scale experimentation was used in this study to investigate and optimize pH for digester stability and biogas productivity.

*2. Test performance of the floor-scale moderate solids system*

Test the stability and scalability of the system. Measure TS, pH, COD, VFA, and biogas production/composition, contributing to efforts to analyze and optimize the system.

*3. Test intermittent mechanical mixing effects in laboratory- scale experiments*

Test a new intermittent mechanical mixing system to keep operational costs low, while ensuring efficient transfer of organic material for the active microbial biomass, release of gas bubbles trapped in the medium, and prevention of sedimentation of denser particulate material.

*4. Develop, calibrate, and validate a process model for the moderate solids anaerobic digester system. Use the model to generate operational insights relevant to pilot scale testing.*

Model calibration is necessary to estimate the most sensitive process kinetics. Model validation is necessary to simulate the new process, predict the behavior of the reactor systems, save experimental time and keep the process development within the time frame of the project.

*5. Test nutrient recovery capabilities from effluent from floor-scale moderate solids system, using previously developed nutrient removal system.*

Effluent from the floor scale system will be treated with a laboratory scale integrated, multifunction ammonia removal and nutrient recovery system (IMARNRS) that has been developed in the Biological Systems Engineering Department of Washington State University, to test nutrient recovery capabilities of the integrated system.

*6. Disseminate results through appropriate scientific publications*

Publish scientific publications, and provide presentations at conferences.

## Section 3: Project approach

### The design concept of the floor scale system, and preliminary optimization for the floor-scale system using laboratory-scale experimentation

The flow chart for the floor-scale moderate solids anaerobic digester system that was built for this project is shown Figure 41. The food waste is fed into the solids digester (20 gallons) (1). Solids and liquids are naturally separated in this tank. Trapped biogas and low specific gravity of food and green waste float the solids. The liquid containing less than 1% TS was pumped into the buffer tank (56 gallons) (3) and then into the UASB seed reactor (two reactors totaling 56 gallons) (2). Effluent was recycled from the UASB reactor to the moderate solids digester in order to provide mixing, pH control, and seeding of anaerobic microorganisms.

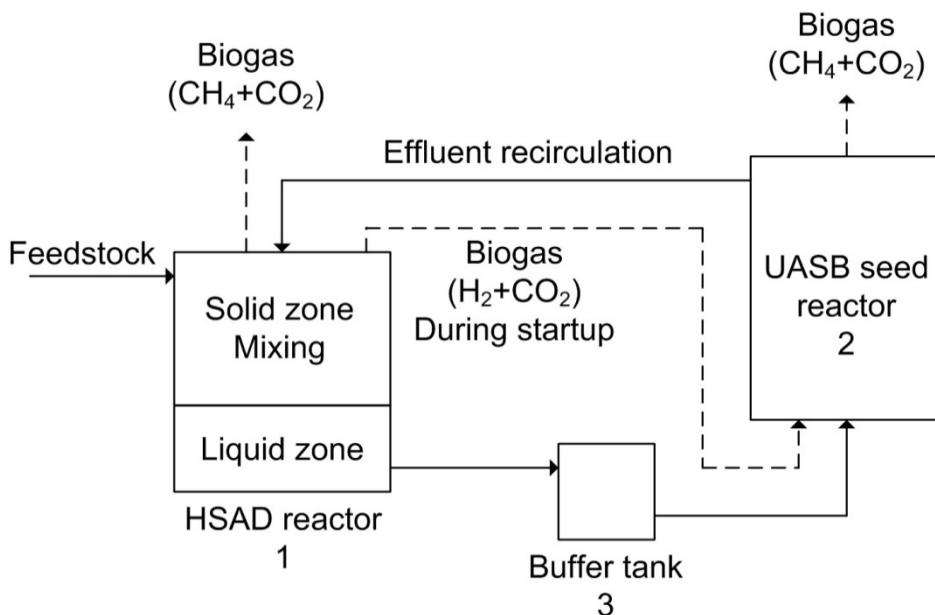


Figure 41. Flow chart of floor scale moderate dual reactor design.

Our use of the UASB seed reactor as a second reactor is based on a long history of use in anaerobic digestion. UASB reactors operate at a relatively high rate compared to other types of digesters (and thus are quite efficient for liquids), and have historically been used most widely for domestic wastewater treatment (Lew et al., 2011). And as the number of anaerobic reactors in the world have increased rapidly in recent years, UASB and EGSB (expanded granular sludge bed) technologies have been used in about 72% of these systems (Anh, 2011). There are a number of operational reasons that UASB reactors have been widely adopted. In up-flow anaerobic systems, it has been observed that bacteria can naturally form aggregate granules under certain conditions (Dolfing, 1986). These dense aggregates have good settling properties and are not susceptible to washout from the system under practical reactor conditions. Retention of granular sludge within the UASB reactor enables good treatment performance at high organic loading rates. Natural turbulence caused by the influent flow and the biogas production provides good wastewater-biomass contact in UASB systems (Seghezzo et al., 1998). Higher organic

loads (in the range of 40 kg COD/m<sup>3</sup>/d) can be applied in UASB systems than in aerobic processes (Kato et al., 1994). Therefore, less reactor volume and space is required while, at the same time, high-grade energy is produced as biogas.

Our design is intended to improve upon the major drawbacks of the three industry standard approaches used for treating moderate- to high-solids feedstocks: (1) wet systems utilizing intensive solids mixing or recycle; (2) dry systems that experience low kinetics and potential product inhibition due to removal of mixing mechanisms; and (3) two-stage acidic and methanogenic separated digestion that require numerous reactors and may suffer from process complexity and/or instability. The dual reactors will minimize mixing expenses by pumping liquid as opposed to solids to achieve mass transfer and desired kinetics. It will also provide for rapid volatile fatty acid (VFA) removal to a liquid reactor reducing concerns regarding VFA and pH inhibition. By operating under an effluent recycle protocol that maintains neutral pH in both reactors, a more stable system is maintained, producing a biogas with high methane content (up to 60-65%).

Another team in the U.S. has been working on similar reactor configurations. However, there are major differences between the two systems in the objective and design of the individual reactors, which collectively make the other system, designed and patented by UC Davis (Zhu et al., 2010), more appropriate for highly cellulosic feedstocks such as rice straw. The system experiences low overall methane production rates when applied to easily degradable waste such as the organic fraction of MSW or animal waste due to low levels of methanogenic bacteria and inappropriate pH in the first step, and low production in the second step. In addition, recycled flows need to be maintained at low rates to avoid washout of the artificial media in the second bio-gasification unit (a continuous anaerobic mixed biofilm reactor), making the system less productive.

In contrast, the system under development by our team is aimed at processing highly degradable wastes, such as animal wastes or the organic fraction of municipal solid waste. Our floor scale design is similar to the laboratory scale system previously tested by our team (Zaher et al., 2009), but with three significant improvements. First, a UASB reactor has been used as the second reactor, improving biogas productivity.

Second, a new intermittent mixing strategy has proven sufficient to break up biowaste clumps (thick, aggregated clumps of food wastes) if the system utilizes two additional passive mixing forces: a design that sprays the effluent from UASB seed reactor into the high solids digester at the top inlet, and natural mixing arising from the force of biogas bubbles rising to the top of the high solids digester (see Objective 3). While this type of strategy has been used previously in wastewater treatment plants, it has not previously been successfully applied for a moderate-solids system such as the one proposed here.

Third, the design took advantage of natural liquid-solid separation processes that were found to occur within the high solids digester. This design improvement was not originally planned, but once discovered, was adopted because it reduces system complexity and operational costs. Though it is recognized that spraying recycled liquid are also used in wastewater treatment plants, maintaining unclogged digesters is substantially more challenging for a moderate solids system.

Optimization of the system conditions was carried out within this project at the laboratory scale prior to testing of the floor scale system, because of the lower cost of optimization experiments at this scale.

### **Testing of floor scale moderate solids dual reactor system**

The floor scale system was built by Andgar Corporation and operated in a hot room maintained at 35 °C (Aquaculture Lab at Washington State University). The UASB seed reactors were built two times larger than the high solids digester. Due to the height limit in the hot room, two UASB seed reactors were built.

Food waste, collected from the WSU cafeteria, was selected as a feedstock, representative of the types of waste that might be collected in a municipal food waste collection program. Anaerobic granular inoculants were taken from Penford Corporation, Richland, WA, a plant that utilizes a UASB digester to treat starch. As with the feedstock, this inoculant was chosen because it was representative of one that might be used in start-up of a food waste system. The floor scale system was tested using the setup shown schematically in Figure 42. Startup of the UASB seed reactor was initiated first instead of the high solids digester. Anaerobic granular inocula were poured into each of two UASB seed reactors. Each gallon of inoculation contained about 2 kg chemical oxygen demand (COD)/m<sup>3</sup> of methanogens.



Figure 42. The floor scale moderate solids anaerobic digestion system.

Food waste often has low pH (less than 5.5) even after storage at low temperature. Therefore, the 30% TS cafeteria food waste was diluted during start-up to less than 5% TS in the high solids digester by the effluent from the UASB seed reactors. The TS feed to the high solids digester

was then gradually increased to 10-20% TS after stabilization for 2-3 days. The recycling rate of effluent from the seed reactor to the high solids digester was maintained at 51 gal/day (relative to a high solids digester volume of 20 gallons, and two UASB reactors totaling 56 gallons in volume), achieving HRTs of about 1 day. This system was tested as a batch process and in continuous feed operating mode. When operated in batch mode, solid food waste was fed once into the high solids digester, after which the whole system maintained biogas production for approximately one month. The system was also operated continuously, at solid waste feed rates of 7.3 kg VS/m<sup>3</sup> per day (calculated from experimental data and using kinetics from the ADM1 model). Performance of both the solids digester and UASB seed reactors was assessed through gas measurement devices operated under atmospheric pressure (Wet Tip Gas Meters, wettipgasmeter.com), which were set as 100 ml at a time to calculate the biogas flow.

### **Testing of intermittent mechanical mixing**

The advantages of mixing have been demonstrated by several researchers (Karim et al., 2005; Buffiere et al., 1998; Kalia and Singh, 2001). Hoffmann et al. (2008) asserted that during anaerobic digestion of sludge, mixing enhances substrate contact with the microbial population, improves the uniformity of pH and temperature conditions, prevents stratification and scum accumulation, facilitates biogas separation from ongoing digestion, and aids in particle size reduction. Seok and Komisar (2003) believed that good mixing promotes the efficient transfer of substrates and heat to the microorganisms, maintains uniformity in other environmental factors and assures effective use of the entire reactor volume by preventing stratification and formation of dead spots and preventing pockets of the VFA from forming.

Table 9 shows a number of parameters of food waste used in this project at different TS levels. The density of the different total solids is 1.002 - 1.012 g/m<sup>3</sup>, and the dry density of food waste is 1.04 - 1.11g/m<sup>3</sup>. The density of food waste is thus very close to that of water and much less than that of manure wastes (1.7 - 2.2 g/m<sup>3</sup>). These values indicate that food waste is easily suspended by water flow. Furthermore, we observed that even 30% TS food waste still has fluidity. It is easy to settle down less than 10% TS food waste in the still water before digestion. The packing limits of food waste in the water is 25% TS while the TS in the moderate solids dual reactor system is 10-15% TS.

Table 9. Observations of food waste with different total solids (TS) percentages.

TS	Density	Average Density	Average TS	VS	Average VS	VS/TS	Dry Density
23.7%				22.48%			
23.4%			23.3%	22.20%			
22.9%		1.012		21.86%	22.18%	95.09%	
17.8%				16.96%			
18.3%			18.1%	17.57%			
18.1%		1.009		17.29%	17.27%	95.51%	
16.0%				15.12%			
15.9%			15.9%	15.12%			
15.8%		1.008		14.94%	15.06%	94.69%	
11.4%				10.80%			
11.6%			11.6%	10.93%			
11.9%		1.006		11.23%	10.98%	94.50%	
11.5%	1.003			10.89%			1.04
11.8%	1.010			11.10%			1.11

TS= total solids; VS= volatile solids

Based on these observations, a simplified mixing strategy of an external pump was used in the floor scale system while a mechanical impeller mixing was tested at the laboratory scale. It was hypothesized that recirculation would be sufficient to promote the retention of microorganisms, needed for high rate digesters (Chynoweth and Isacson, 1987). This mixing system reduces energy consumption and therefore operational costs compared to the previous design using a helical auger.

## Development of the process software model

Process software has been used commonly to support full-scale design, operation, and optimization of anaerobic digesters. Anaerobic Digestion Model No.1 (ADM1) software was developed for this purpose (Batstone et al., 2002; Fedorovich et al., 2003), and has been widely applied because it contains the most complex biochemical reaction kinetics in anaerobic digestion such as hydrolysis, equilibrium, substrate uptake and inhibition. Since ADM1 model was published by IWA task group, it has been tested and proven successful in simulating the anaerobic digestion of several organic wastes such as: industrial wastewaters (Batstone and Keller, 2003); sludge from wastewater treatment plants (Blumensaat and Keller, 2005; Parker, 2005); sewage sludge (Shang et al., 2005); black water from vacuum toilets (Feng et al., 2006); and olive mill solid waste (Kalfas et al., 2006).

ADM1 is a structured model that reflects the major processes that are involved in the conversion of complex organic substrates into methane and carbon dioxide and inert byproducts. Previous work by our group (Zaher et al., 2009) included the development of a general co-digestion model (GISCOD) using the Matlab-Simulink® implementation of the International Water Association (IWA) ADM1 to study the anaerobic degradation and methane production from mixed food waste feedstocks are used (e.g. food waste and manure). GISCOD applies an

advanced transformer model procedure to estimate the composition and study the hydrolysis of each waste separately.

Since the new digester design is appropriate for digesting food waste alone, a new process model was needed to assist in the scale-up to the floor-scale moderate solids system. The model concept for this project was developed from the IWA ADM1 using the schematic shown in Figure 43. The two-stage AD system with a high solids digester and a liquid digester was modeled with the new ADM1 code. The effluent was re-circulated from liquid digester to high solids digester. The high solids digester was modeled as a combination of continuous stirred-tank reactor (CSTR) and dispersed plug flow reactor (PFR). The liquid digester was assumed to be a dispersed PFR where UASB, a high rate digester, was used to retain more methanogens and maintain pH throughout the system in the range of 6.0-8.0. The model was verified by the data from the floor-scale experiments and parameters in the kinetics models were corrected as needed.

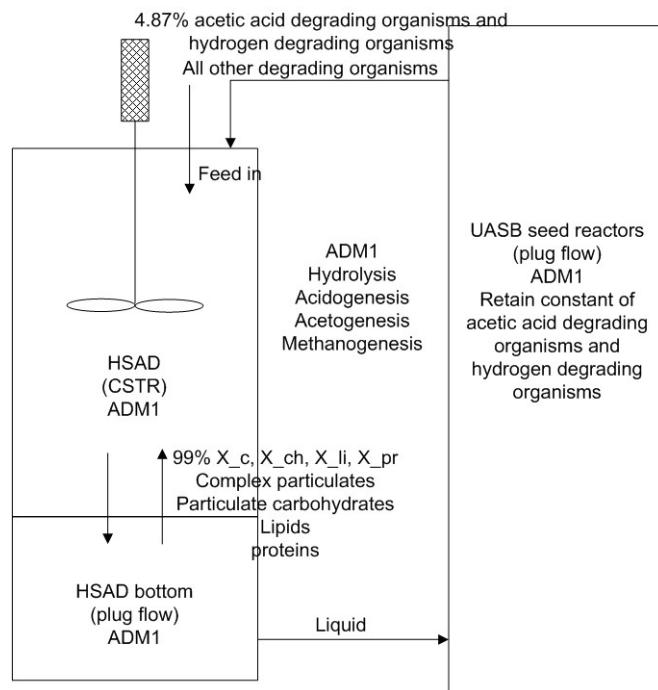


Figure 43. Process model of the floor scale moderate solids anaerobic digestion system.

## Nutrient recovery

Effluent from the floor-scale system was experimentally treated in a laboratory-scale integrated, multifunction ammonia removal and nutrient recovery system (IMARNRS), that has previously been developed, pilot-tested and now commercially demonstrated on a Washington state dairy for AD manure effluent treatment (Figure 44; Jiang et al., 2009). Through IMARNRS, high concentrations of nutrients, a potential pollutant if field-applied, are converted to value-added byproducts, helping to offset the costs of implementing nutrient recovery.



Figure 44. IMARNRS pilot testing facilities.

The IMARNRS process (Figure 45) begins with enhanced solids settling of the AD effluent through dilution and/or flocculent addition. Importantly, this settling and removal of the solids also allows for less costly and more efficient pH control in the next step. The supernatant then flows into the second stage settler where lime (or lime with polymer addition such as Polyacrylamide (PAM)) is added to raise the pH to about 10. This promotes precipitation of phosphorus and allows for the production of a slow release phosphorus fertilizer composed of calcium carbonate, calcium phosphate, and organics. The effluent is then sent to a closed-loop stripping tower, where ammonia is absorbed through addition of sulfuric acid to make a concentrated liquid ammonia sulfate for export off the farm. The resulting effluent will still be high in pH; therefore biogas produced from the digester is bubbled through the effluent to reduce its pH to about 8 prior to its return to the manure storage lagoon. This step simultaneously purifies the biogas by absorbing CO<sub>2</sub> and H<sub>2</sub>S. Recycling the low nutrient effluent back through the IMARNRS process can be optionally implemented, and has been shown to aid in the solids settling step (Jiang, 2010) and the ammonia removal process, though it may also add to costs.

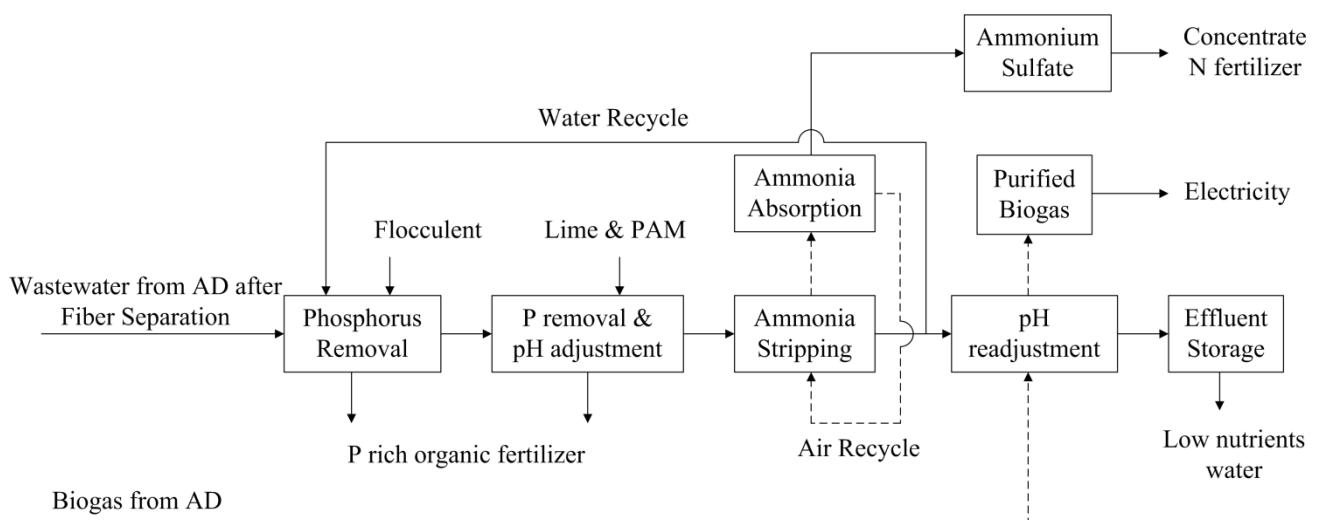


Figure 45. Schematic Diagram of IMARNRS.

Food waste was tested to examine its suitability for the IMARNRS process. The major differences between food waste and manure are summarized in Table 10. Food waste contains higher protein levels than manure and thus contains significant amounts of nitrogen and phosphorus. When mineralized to ammonia and phosphate, these nutrients can have inhibitory effects on the anaerobic digestion process. To avoid operational failure in the absence of nutrient recovery, the organic loading rate (OLR) must then be limited.

Table 10. Characteristics of diluted manure and kitchen waste.

Characteristics	Unit	Diluted Manure Waste	Food Waste
Total Chemical Oxygen Demand (CODt)	(gCOD m-3)	27,217	380,647
Particulate COD (CODp)	(gCOD m-3)	23,550	368,400
Soluble COD (CODs)	(gCOD m-3)	3,667	12,247
Soluble COD without VFA COD(CODs-VFA)	(gCOD m-3)	2,521	3,500
Volatile Fatty Acids (VFA)	(gCOD m-3)	1,146	8,747
Total Carbon (TC)	(gC m-3)	10,064	139,760
Total Organic Carbon (TOC)	(gC m-3)	9,340	139,280
Total Inorganic Carbon (TIC)	(mol HCO3- m-3)	60	40
Total Kheldal Nitrogen (TKN)	(gN m-3)	882	15,300
Total Organic Nitrogen (Norg)	(gN m-3)	598	14,000
Total Ammonia Nitrogen (TAN)	(gN m-3)	284	1,300
Total Phosphorus (TP)	(gP m-3)	219	1,606
Organic Phosphorus (TP-orthoP)	(gP m-3)	187	720
Ortho-Phosphate (orthoP)	(gP m-3)	32	886
Total alkalinity (S cations)	(equ m-3)	60	25
Total Solids (TS)	(g m-3)	20,697	291,000
Fixed Solids (FS)	(g m-3)	5,397	31,000
Total Volatile Solids (TVS)	(g m-3)	15,300	260,000
Carbohydrate	(g m-3)	10,924±428	153,400±11,180
Protein	(g m-3)	4,069±367	85,800±8,320
Lipids	(g m-3)	306±61.2	20,800±2,860

For testing of the IMARNRS process, effluent was continuously recycled for two weeks without any dilution in the floor-scale moderate solids dual reactor system. Then a 5 L effluent sample was taken from the buffer tank. The lab-scale nutrient recovery system shown in Figure 46 was used to remove ammonia and phosphate from the effluent. 400 ml UASB effluent was placed in a 500 ml serum bottle, maintained at 55 °C, and aerated for 24 hours. Samples were taken every two hours during aeration. Alkalinity was measured by titration using 0.2 N H<sub>2</sub>SO<sub>4</sub>. After aeration, AD effluent was settled for 24 hours, and then total phosphorus (of the supernatant) and total ammonia nitrogen (of the supernatant and settlement) was measured. Total phosphorus (TP) was determined according to standard methods (Eaton and Franson, 2005). Total Kjeldahl nitrogen (TKN) was measured using a 2300 Kjeltec Analyzer Unit (Tecator, Perstorp Analytical).



Figure 46. The lab-scale nutrient recovery system.

## **Section 4: Project results**

### **Objective 1: Floor-scale design and build-out of the moderate solids system, and preliminary optimization for the floor-scale system using laboratory-scale experimentation**

Floor-scale build-out of the system was achieved, with the design advantages over the laboratory scale system described above. This system successfully treated moderate (10% - 20%) solids waste for six months with stable operational status.

#### **Fabrication of the floor-scale moderate solids system**

Floor-scale build-out of the moderate solids reactor systems was successfully completed. The completed system was scaled approximately 10 times larger than the previously tested laboratory scale system. The system included one high solids digester (20 gallons), two UASB-seed reactors (56 gallons) and two buffer tanks (56 gallons). The effective reactor space is about 100 gallons. The engineering drawings are shown in Appendix A.

#### **Design improvement through liquid-solid separation**

In a mixed solids digester, VFA must be removed from the reactor quickly to avoid provoking a drop in reactor pH and process failure. The traditional European high solids digestion technologies depend on recycling some of the treated solids to maintain adequate bacterial populations that consume VFA in the digester. Additional reactor volumes and expensive equipment are required to recycle the treated solids.

Our experimental observations suggested that it might be possible to remove liquid only from the reactor, reducing complexity and costs. As can be seen in Figure 47, liquids and solids separate naturally. The left bottle was filled with a mixture of food waste and inocula from the effluent of UASB seed reactors. The right bottle was filled with food waste and water. In the left bottle, biogas production pushed the food waste up to the top level in less than one day. TS of the food waste aggregated in the top level reached as high as 22%. In the right bottle, the food waste was floated up slowly and the separation of liquid and solid has only begun in the second day. This phenomenon can be used to separate solids from liquid in the solids digester, allowing solids to remain in the high solids digester while the liquid with VFA is recycled to the UASB seed reactors.



Figure 47. Natural separation of liquid and solid in a mixture of food waste and inocula from the effluent of UASB seed reactors (left bottle), and food waste and water (right bottle). (Left picture: after one day, Right picture: after two days).

### **Determination of optimal operational conditions for the floor-scale moderate solids system using laboratory scale experimentation**

In the traditional two stage anaerobic digestion system, the hydrolytic/acidogenic process (that breaks complex organics down into simpler organic materials) is separated from the methanogenic process (that generates methane) in order to bring maximum control over the bacterial communities living within the digesters (Chynoweth and Isaacson, 1987). The first digester is often designed as hydrolytic/acidogenic process where H<sub>2</sub> and CO<sub>2</sub> are major gas products (Jagadabhi et al., 2011) and pH is maintained at between 4.5 and 6.0 (Zhang and Zhang, 2002). In contrast, the moderate solids system operates at a pH that allows for methanogen activity (and thus methane production) in the first digester as well as the second.

During optimization experiments, we determined the optimal pH for productivity by measuring VFA concentrations under a variety of pH conditions. These experiments were carried out in a 7.5 L fermentor designed to represent digestion in the high solids digester (BIOFLO 110, New Brunswick Scientific, USA). Figure 48 shows the change of total VFA with time at the different pH. The highest VFA productivity (and therefore presumably subsequent methane production) was obtained in the range of pH 6 to 7. This is consistent with Chynoweth and Isaacson (1987), who found that optimum pH for anaerobic digestion, is in the range of 6.0 - 8.0 with pH levels that deviate significantly from this range indicating potential toxicity and digester failure. At pH levels below 6.0, the acidic conditions produced can become toxic to methane bacteria. These results emphasize the importance of recycling adequate effluent from the UASB seed reactors to maintain pH in the high solids digester. VFA compositions (Figure 49) and TS reduction (Figure 50) are also dependent on pH. These results further confirm that the optimum range for hydrolysis and acidogenesis processes in the high solids digester is 6.0 - 8.0.

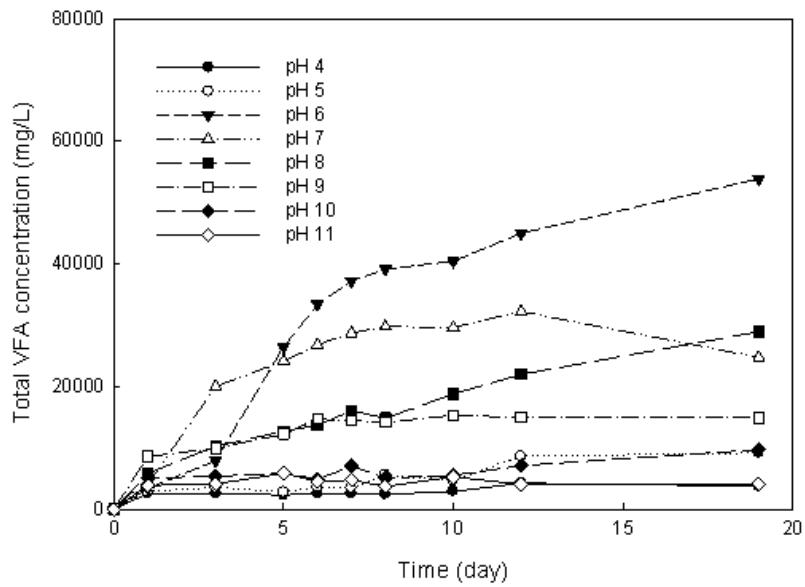


Figure 48. Change of total VFA with time at the different pH.

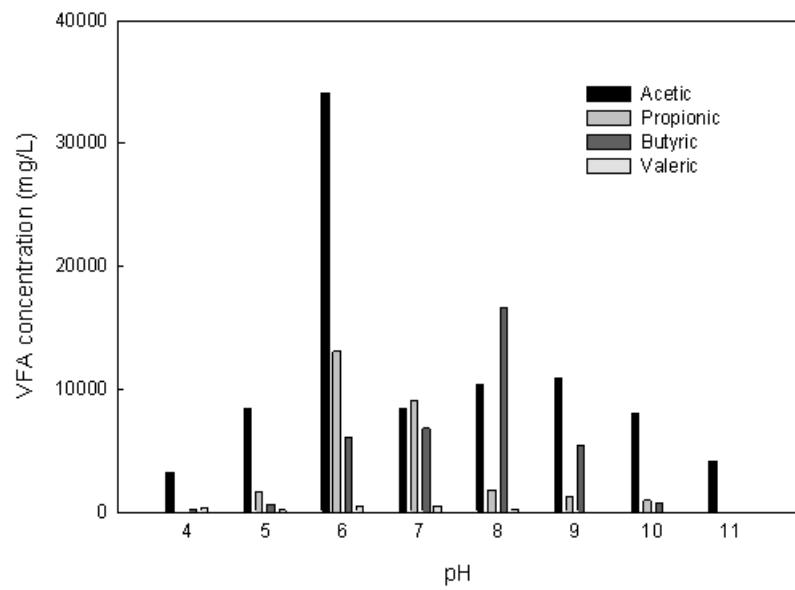


Figure 49. Change of VFA concentrations with pH.

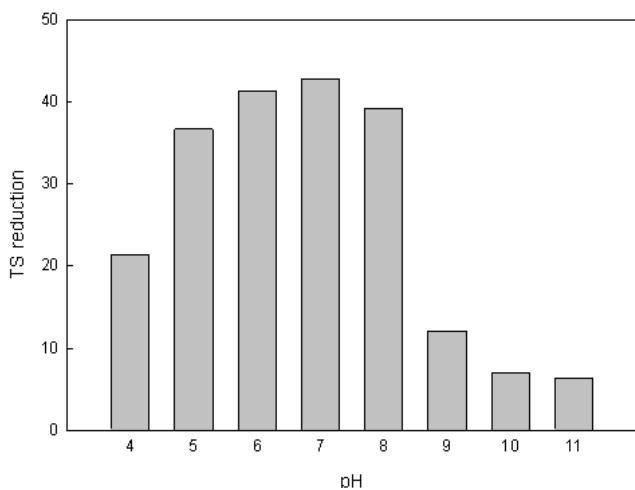


Figure 50. Change of TS reductions with pH.

### **Comparison of floor-scale high solids anaerobic digester system to existing systems**

In order to understand the advantage of our new concept design, a comparison was made between the new HSAD and an anaerobic phased solids (APS) digester process recently developed by Dr. Ruihong Zhang in the Biological and Agricultural Engineering Department at the University of California, Davis (Zhang and Zhang, 2002), that incorporates a new, high rate bioconversion technology. Table 11 compares the system characteristics of these two systems. There are different units of pretreatment, liquid-solid separation and system equipment in these two processes. The simplicity of the HSAD system compared to the APS system could potentially save money in fixed asset investments. The pH of the high solids digester in the new HSAD can be maintained at the range of 6.0 - 8.0, the optimum range for methanogenic bacteria. This leads to the production of more bio-methane. Furthermore, the UASB provides effluent to the high solids digester for mixing, pH control, and seeding of anaerobic microorganisms. This allows higher up-flow velocities in the range of 4 - 10 meters per hour and most of methanogenic bacteria can be retained in the UASB avoiding washout. The hydraulic retention times (HRTs) range from 6 to 24 hours (Seghezzo et al., 1998). The organic volumetric loading rate for VFA is up to 15000 – 24000 mg/L at the temperature of 35 °C (Lettinga and Pol, 1991). This makes it possible to recycle enough effluent to stabilize and accelerate the process of the high solids digester.

Table 11. Comparison of new HSAD and APS-digester in system characteristics.

	New HSAD	APS-digester (Zhu et al., 2010; Zhang and Zhang, 2002)
Pretreatment	Separated in HSAD reactor. Heavy materials will settle down and directly discharge.	Using rotary drum reactor to separate plastic, metals, glass and stone etc.
High solids reactor	pH: 6.0-8.0. Produce methane by recirculation of enough effluent	pH: 4.5-6.5. Produce CO <sub>2</sub> and H <sub>2</sub>
Liquid-solid separation	Biogas floating and gravity flow	Positive drive pump inside the hydrolysis reactor and screen filter system
Bio-gasification reactor	Upward-flow anaerobic sludge blanket (UASB)	Continuous anaerobic mixed biofilm reactor
System	More simple, including high solids digester, buffer tank, and UASB	More complex including 6 reactors

In addition, the HSAD system produces primarily methane, already easily used in the existing natural gas system. Although hydrogen, produced by the APS digester, is a promising clean energy, so far commercialization of hydrogen has been prevented by the fact that it has been difficult to produce high purity hydrogen through anaerobic digestion.

## Objective 2: Floor-scale testing of the moderate solids system

### Floor-scale testing of the high solids digester

In the high solids digester, solids retention times (SRTs) are different from HRTs. Under batch mode operation, SRTs were 7 days, while HRTs were less than one day, to allow for separation of liquid and solid in the high solids digester. Under continuous mode operation, SRTs were 30 days at a solids loading rate of 5.0 - 15.0 kg VS/m<sup>3</sup> day, while HRTs were again less than 1 day.

Figure 51 shows the effect of recirculation ratio on the pH of the high solids digester. At the beginning, the pH of the food waste fed in the high solids digester was 4.5 - 5.5 while the pH of effluent was 7.9. The pH decreased with time and with a decrease in the rate of effluent recirculation. The results suggest that HRTs less than 1 day could maintain pH over 6.0 even at the low recirculation ratio. Low recirculation ratio also means the volume of the UASB seed reactor can be reduced, reducing costs. Alternatively, the UASB could provide more capacity for additional solids reactor volume. Optimum sizing at the commercial scale will likely need to be modeled in order to assess main reactor volume and UASB reactor volume.

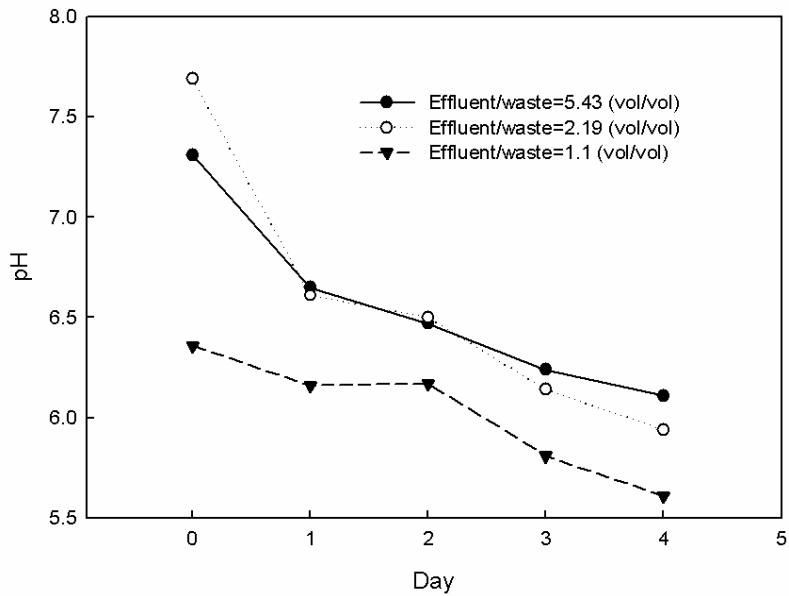


Figure 51. Effect of recirculation ratio on the pH of the high solids digester.

Figure 52 shows the daily biogas production rate in the high solids digester during the batch study, which averaged about 10 L/day. The highest biogas production was 40 L/day. The pH of the high solids digester was 6.5 - 7.5, with stable operation on a daily and monthly basis. No trend towards acidification occurred while effluent was recycled into the high solids digester. On the first day, more effluent was needed to remove the VFA and maintain pH in the high solids digester because long-term storage of food waste had resulted in degradation, producing more VFA even at the relatively low temperature used.

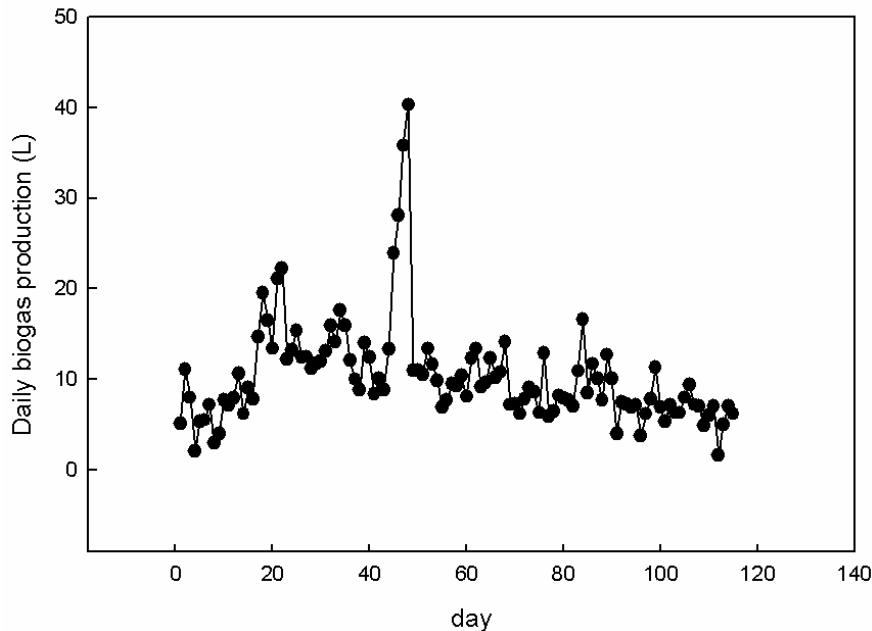


Figure 52. Daily biogas productions in the high solids digester (experimental data).

#### Floor-scale testing of the upflow anaerobic sludge blanket seed reactors

Figure 53 shows the exponential decay of COD and VFA over time in the experimental UASB seed reactors. The VFA removal in the first day is 38.3% and the total VFA removal is 63.4%. The COD removal in the first day is 32.0% and the total COD removal is 56.1%. The results indicate that the highest rate of COD and VFA removal by the UASB seed reactor occurs in the first day, leading to a recommendation that hydraulic retention times of 1 day be used in order to improve the efficiency of the UASB seed reactors.

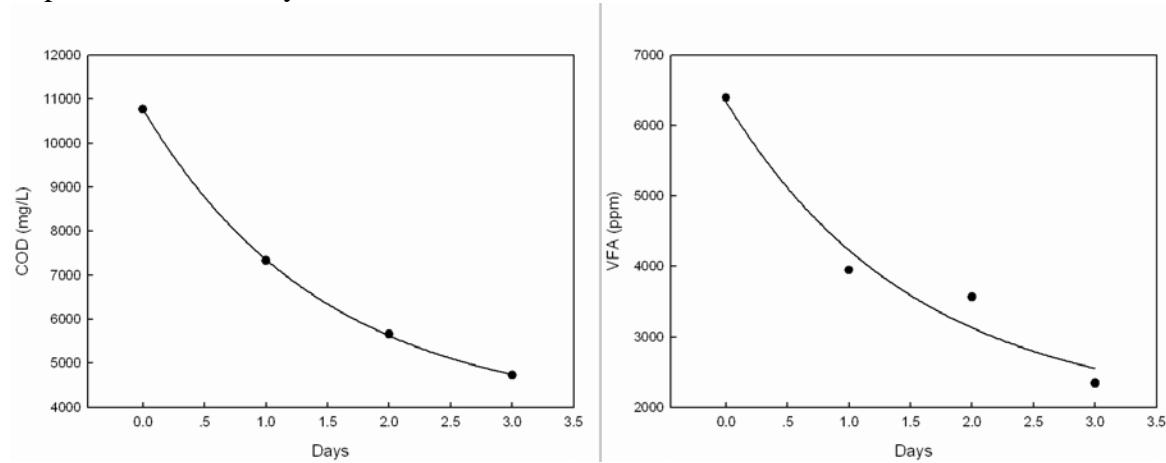


Figure 53. Change of COD and VFA over time in the UASB seed reactors.

Figure 54 shows the cumulative biogas production in the UASB seed reactors when leachate was fed from the high solids digester once (10 gal/week) and then recirculated within the solids reactor (all runs). Effluent was also recirculated between the UASB seed reactors and the buffer tank, at

varying rates. The results show that a higher recirculation rate enhanced the cumulative biogas production rate until day 2. After day two, it is likely that the lower levels of biogas production (seen in all experimental runs) were due to the depletion of COD and VFA (Figure 13).

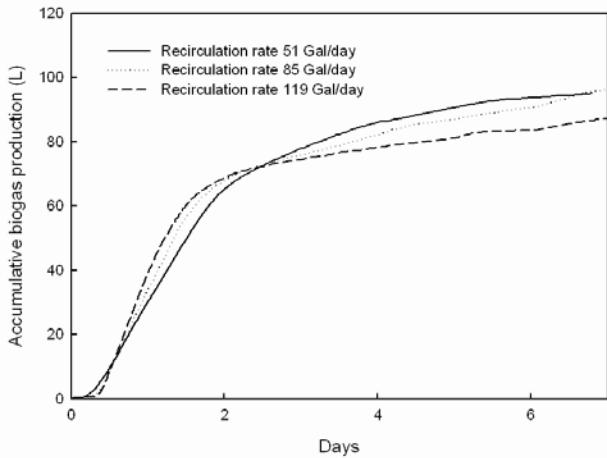


Figure 54. Cumulative biogas production at a feeding rate of 10 gal/week in the UASB seed reactors.

Figure 55 shows the cumulative biogas production in the UASB seed reactors when leachate from the high solids digester was continuously fed to the UASB, at a rate of 10 gal/day. The cumulative biogas production showed a linear increase, indicating that the short hydraulic retention times of less than 1 day in UASB seed reactors can be used to improve biogas production. Within this range, mixing enhanced mass transfer to accelerate the release of biogas.

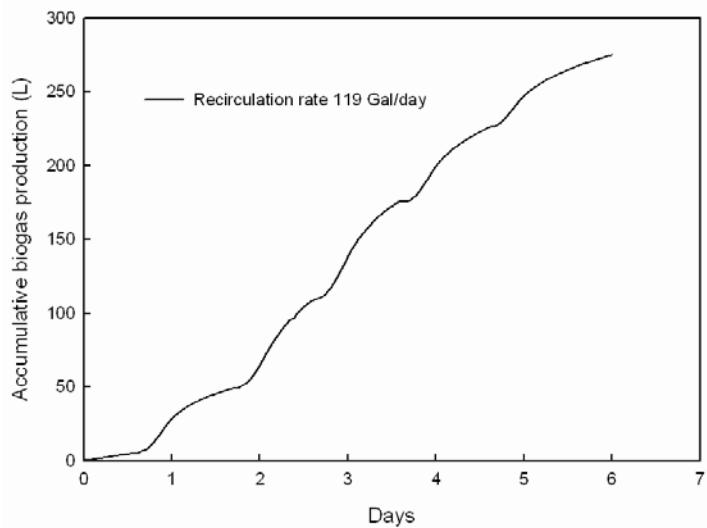


Figure 55. Cumulative biogas production at a feeding rate of 10 gal/day in the UASB seed reactors.

Figure 56 shows the daily biogas production in the UASB seed reactors, which fluctuated over time. Average biogas production was about 40 L/day. The highest biogas production was 91 L/day. The pH of UASB seed reactors was about 7.4 - 7.6. The CH<sub>4</sub> content of the biogas was 63.7%. The biogas production in the UASB seed reactors was about 4 times higher than that in the high solids digester because there were more methanogenic bacteria in the UASB seed reactors. (More methanogenic bacteria persist in the UASB reactor, but recycled material also provides methanogenic bacteria and alkalinity to the high solids digester, helping to control pH and allowing for methane production in the high solids digester.)

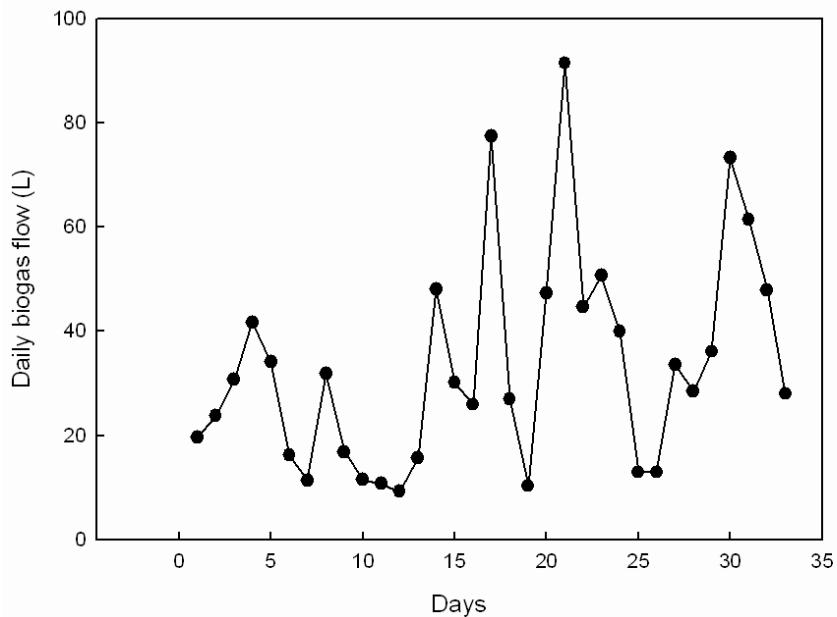


Figure 56. Daily biogas production in the UASB seed reactors.

### Conclusions of floor-scale testing of the high solids anaerobic digestion system

The floor scale system was operated stably for a six month period. During this time, COD, VFA, biogas production, methane content, TS, and VS were monitored. Table 12 shows the higher productivity in terms of methane production rates, methane yields, and methane content, of new HSAD system compared to other existing systems, based on the laboratory scale optimization. Methane production rate, methane yield, and methane content are all higher using our HSAD system than other existing systems.

Table 12. Comparison of new HSAD and existing systems in productivity.

Digester System	New HSAD	APS	Batch	SEBAC	Conventional HSAD system with solid waste recycle	HSAD using Kompogas system
Total solid loading (g/L)	100	100	100	75	40	26
Biogas production rate (m <sup>3</sup> /m <sup>3</sup> /day)	3.2				3	2.8
Methane production rate (m <sup>3</sup> /m <sup>3</sup> /day)	2.0				1.52	
Methane yield (L/g VS)	0.29	0.42	0.05	0.35		
Methane content in biogas (%)	63.9	50.6	27.72	40.78		
Total solids reduction (%)	46.42	40.67	5.33	36.21		
Volatile solids reduction (%)	48.63	49.14	8.01	41.08		
COD removal (%)	47.33				45.38	

Note: SEBAC, sequential batch anaerobic composition

In conventional digesters, food waste digested alone readily degrades but tends to acidify quickly, inhibiting digestion. In contrast, our floor-scale dual reactor system treated moderate (10% - 20%) solids waste for six months with stable operational status under batch feed operation. Operation and performance evaluation were ongoing through July 2011. The stable operation in spite of the fluctuation of organic loading rate demonstrates a process stability that is encouraging for future efforts to advance the system to pilot and commercial scale.

It is also worth noting that our research suggests the possibility that the new HSAD system could be applied at lower psychrophilic temperatures (rather than the more common mesophilic temperatures), as the food waste was observed to be hydrolyzed at low temperatures (4 °C) during storage. This possibility is supported by other research on food waste (Lew et al., 2004), though further work will be needed to explore this issue. This is in contrast to researchers working with other feedstocks, who have in general reported declining performance at lower temperatures. At the mesophilic temperatures of 30 – 38 °C COD removals above 70% have been reported by several authors (Lew et al., 2002; Kalogo and Verstraete, 2000). At lower ambient temperatures, overall performance deteriorated with COD removals of 65% at 20 °C and 55 -65% at 13 – 17 °C (Lew et al., 2004; Al-Jamal and Mahmoud, 2009). At low temperatures, the performance of intensive anaerobic systems is often highly limited by the hydrolysis of suspended solids and/or the decrease in the organic matter degradation, which accumulates in the sludge bed. If our system is applicable at lower temperatures, it could particularly benefit cooler climates such as Washington.

### Objective 3: Effect of mechanical mixing in the laboratory-scale experiments

While a simplified mixing strategy of an external pump was used in the floor scale system, future scale-up will almost certainly need a more intensive mixing system. A mechanical impeller mixing was tested at the laboratory scale. It was hypothesized that the mechanical impeller mixing might be sufficient for future scale-up work, reducing energy consumption compared to the previously proposed design using a helical auger.

Figure 57 shows the effect of mixing in laboratory scale studies on the VFA and pH in contour and surface plots. The plots show the VFA increases with increased mixing level and time. The pH decreases with an increase of mixing level, but increases with an increase of time.

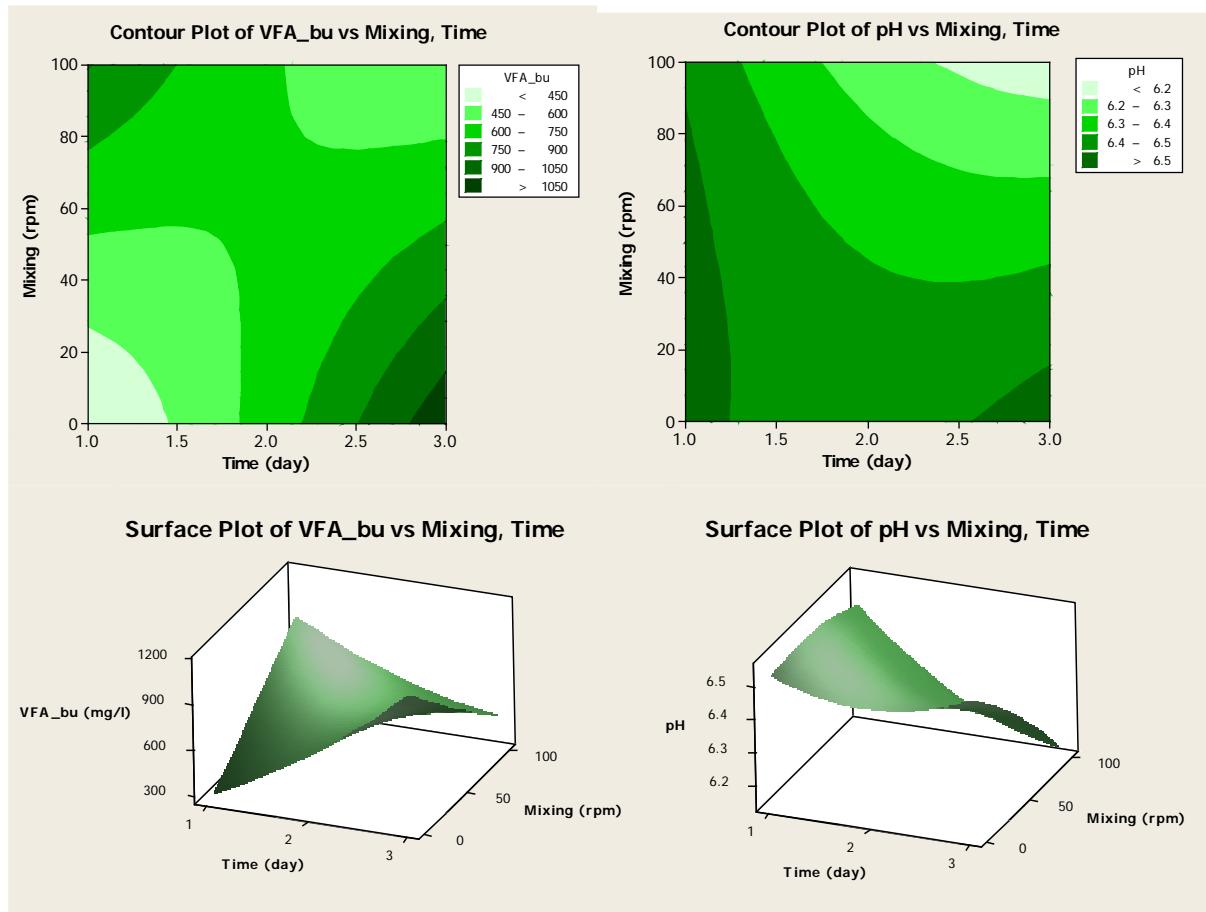


Figure 57. Effect of mixing on VFA and pH.

These results confirm others' work reporting that propionate and acetate accumulated in high levels in continuously mixed digesters that were subjected to aggressive startups and overloading (Stroot et al., 2001; McMahon et al., 2001). While acetate was eventually consumed, propionate persisted throughout system operation. When the mixing level was reduced, propionate was degraded and digester operation was stabilized, indicating that adjustment of the mixing level could serve as an effective operational tool for stabilizing digesters. Vavilin and Angelidaki (2005) believed a vigorous mixing is not suitable in a system where methanogenesis is the rate limiting step, because imperfect mixing conditions create refuge areas where methanogenic bacteria are protected from rapid acidogenesis. During the startup period it is better to avoid a vigorous mixing, which would suppress methanogenic population centers and therefore growth and propagation over the reactor volume. If hydrolysis of large molecules becomes the rate- limiting step of the overall anaerobic process, when methanogenic biomass is sufficient in the reactor, mixing may be beneficial for maximizing the degree of hydrolysis.

In applying these lessons to future pilot-scale development of our two stage anaerobic digester, we hypothesize that continuous and vigorous mixing could be used if the UASB seed reactor was big enough to consume VFA and maintain pH by recycling effluent to the high solids digester. Considering the separation of liquid and solid and the power consumption of agitators, an intermittent mixing with enough intensity to break up the biowaste clumps (thick, aggregated clumps of food waste which naturally form) is recommended.

## **Objective 4: Development of a mathematical model for the moderate solids anaerobic digester system and operational insights for scale-up**

### **Development of a mathematical model**

The process model developed by our team, and based on the ADM1 model (Batstone et al., 2002; Fedorovich, 2003), includes hydrolysis of complex solids into inert substances, carbohydrates, proteins and fats during stable operation of the anaerobic digestion process. The products of this disintegration are hydrolyzed to sugars, amino acids and long chain fatty acids (LCFA). Carbohydrates and proteins are fermented to produce volatile organic acids (acidogenesis) and molecular hydrogen. LCFAs are oxidized anaerobically to produce acetate and molecular hydrogen. Propionate, butyrate and valerate are converted to acetate (acetogenesis) and molecular hydrogen. Methane is produced by both cleavage of acetate to methane (aceticlastic methanogenesis) and reduction of carbon dioxide by molecular hydrogen to produce methane (hydrogenotrophic methanogenesis).

Two main types of reactions were included in the model: biochemical reactions and physicochemical reactions. All extracellular biochemical reactions, which are catalyzed by enzymes acting on the pool of biologically available organic material, were assumed to be first order. This is a simplification based on empirical data, reflecting the cumulative effect of a multi-step process. Substrate uptake Monod-type kinetics were used as the basis for intracellular biochemical reactions. Biomass growth is implicit in substrate uptake. Inhibition functions included pH (all groups), hydrogen (acetogenic groups) and free ammonia (aceticlastic methanogens). pH inhibition was implemented as one of two empirical equations, while hydrogen and free ammonia inhibition were represented by non-competitive functions. The other uptake-regulating functions were secondary Monod kinetics for inorganic nitrogen (ammonia and ammonium), to prevent growth when nitrogen is limited, and competitive uptake of butyrate and valerate by the single microbial group that utilizes these two organic acids. Physicochemical reactions are not biologically mediated by microorganisms and encompass ion association/dissociation and gas-liquid transfer. Dissociation/association processes are very rapid compared to other reactions (especially biochemical); they are often referred to as equilibrium processes. They were thus described by algebraic (rather than differential) equations. Liquid-gas transfer was described using dynamic gas transfer equations, based on two-film theory.

Table 13 shows the initial conditions used in the mathematic model for the moderate solids anaerobic digester system. The TS of the feedstock was 15.4%. Figure 58 shows daily biogas production as predicted by the model, compared to experimental data from the UASB seed reactors, under a feed rate of leachate from the high solids digester of 10 gal/week. The model shows good agreement with the measurement of biogas production in the floor scale UASB seed reactors.

Table 13. Conditions for the mathematic model for the moderate solids anaerobic digester system.

Characteristic	Unit	Solid food waste
Total solids (TS)	g/l	154
Volatile solids (VS)	g/l	118
Particulate COD (CODp)	gCOD/l	182
Soluble COD (CODs)	gCOD/l	18.8
Total organic carbon (TOC)	gC/l	53.3
Inorganic carbon (IC)	mole/l	10.7
Total Kjeldahl nitrogen (TKN)	gN/l	5.3
Total ammonium (TAN)	gN/l	1.3
Total phosphorus (TP)	gP/l	4.6
Total ortho-phosphorus (OP)	gP/l	0.9
Volatile fatty acids (VFA)	gCOD/l	2.3
pH		5.47

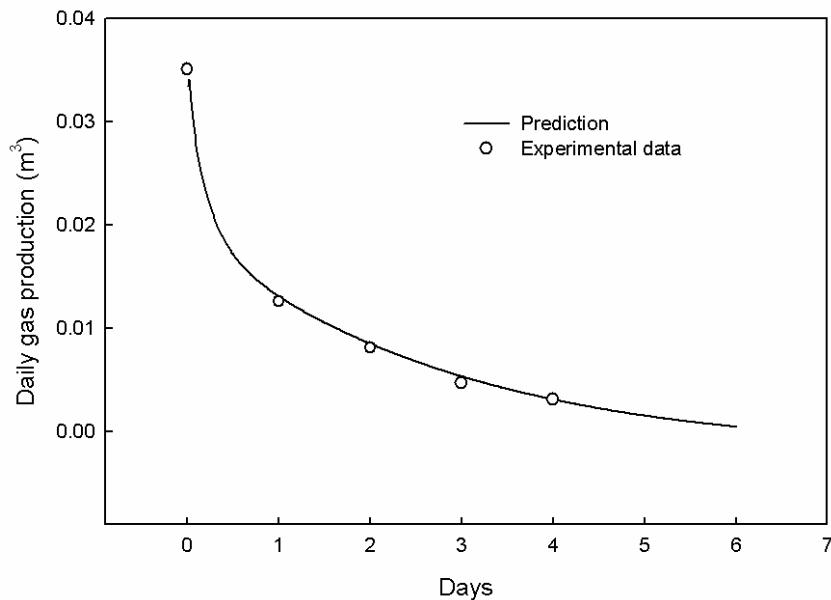


Figure 58. Comparison of daily biogas production between prediction and experimental data from UASB seed reactors.

Figure 59 shows a comparison of predicted and experimental VFA levels from the UASB seed reactors. The major composition of VFA includes acetic acids, propionic acids, butyric acids and valeric acids. Static headspace gas chromatography was used to analyze VFA composition. Compared with the other VFA compositions, the consumption rate of propionic acids was relatively slow. The concentration of acetic and propionic acids are critical factors that regulate anaerobic digestion processes, since oxidation of propionic acid to acetic acid is the slowest among the VFA (Sonakya et al., 2007; Dohanyos et al., 1985; Inanc

et al., 1999). The model shows good agreement the measurement of acetic acid, propionic acid, and butyric acid in the floor-scale UASB seed reactors. However, the calculated results of valeric acids do not fit well with the measurements. This is because butyric acid and valeric acid share the same equation for degrading organism in the ADM1 model, and are proportionally related, though our experimental data does not indicate a proportional relationship. However, for the engineering purposes that these results are being used for (and will be used for in the future), these results are close enough to predict the most important features of the digestion process.

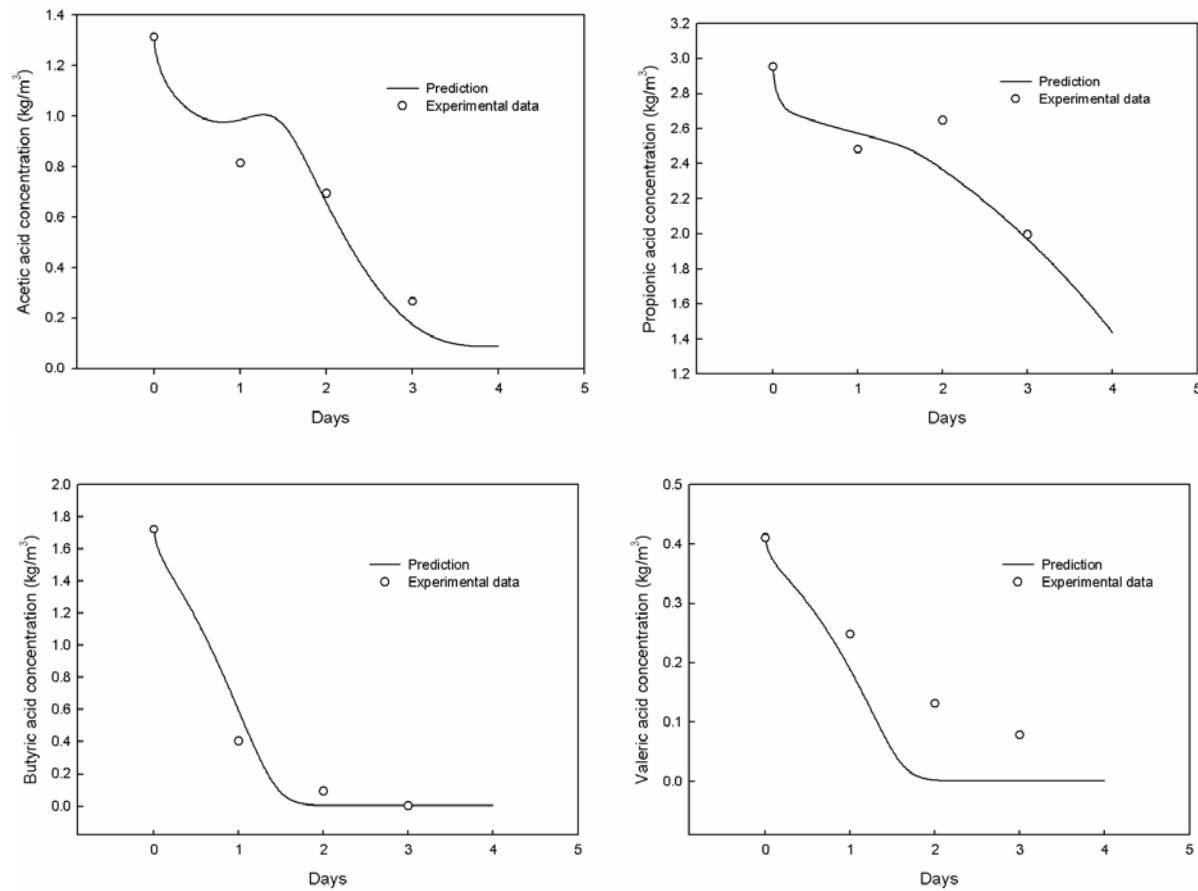


Figure 59. Comparison of VFA between prediction and experimental data from the UASB reactor.

### Operational insights from mathematical model

Based on the above comparisons of model predictions and experimental results, the model developed in this project provides a good description for a moderate solids design. We were thus able to use the model to run virtual experiments that would be difficult to run experimentally. These types of results can be used to focus future experimental testing and verification. Since the effluent recycled from the UASB seed reactors to the high solids digester is critical for stabilizing the moderate solids dual reactor system and enhancing methane productivity, the new HSAD model was used to understand the quantitative impacts of continuous change of recirculation rate, which is difficult to obtain from experiments.

Figure 60 shows the predicted effect of recirculation rate on pH. When the recirculation rate is lower (Figure 60A), the pH in the high solids digester goes down significantly over the first two days due to a large amount VFA produced in the fresh feedstock of the high solids food waste. After the VFA in the high solids digester is carried to the UASB seed reactors by effluent, the pH goes back to 7.5 - 8.0. The pH in the UASB seed reactor is maintained at 7.5 - 8.0 and most of VFA is consumed in this reactor. When the recirculation rate is higher, as shown in Figure 60C, the VFA can be carried to the UASB seed reactors by effluent more quickly. Therefore, the pH in the high solids digester is maintained at 7.5 - 8.0.

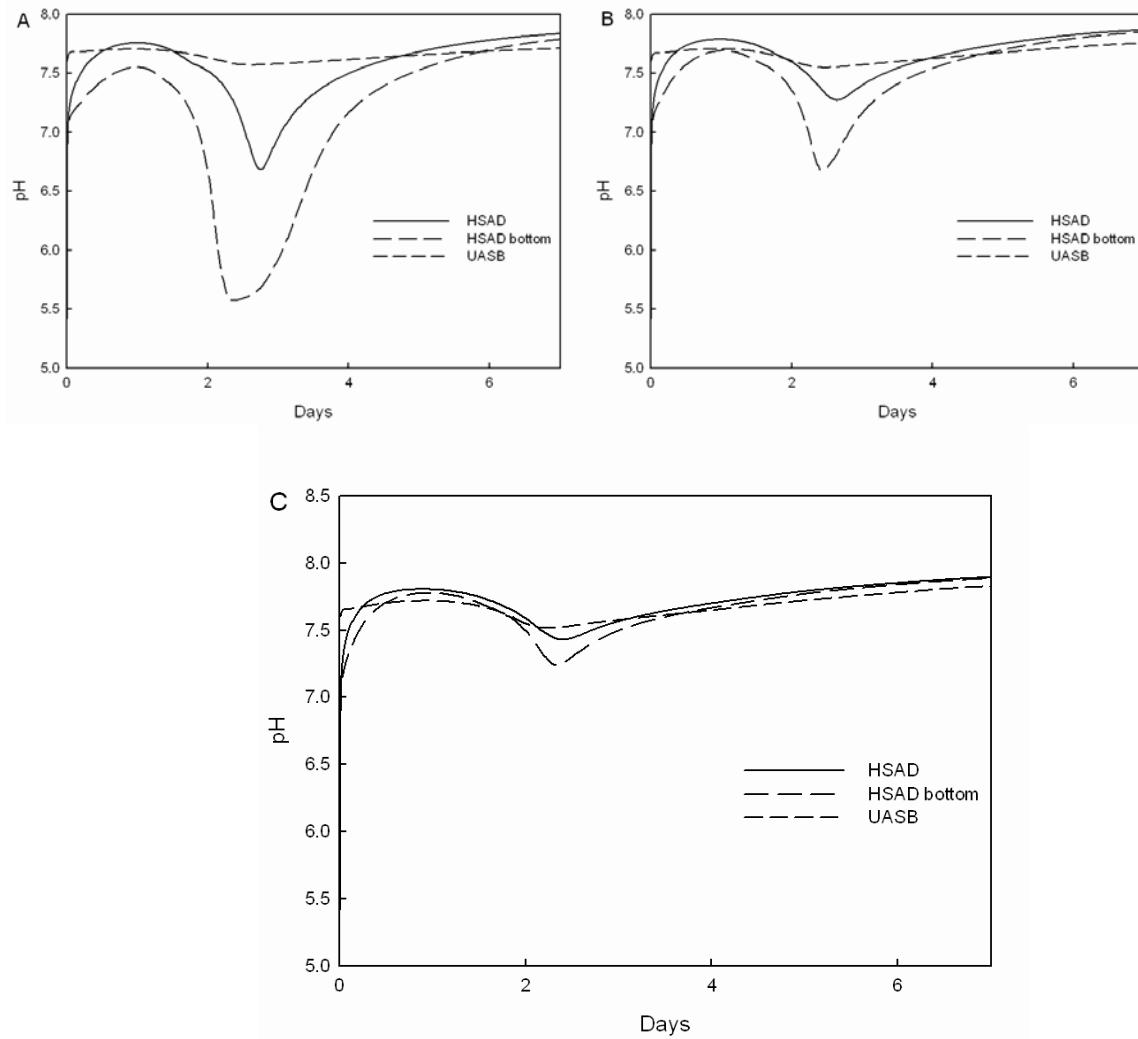


Figure 60. Predicted effect of recirculation rate on pH (A) 0.09 m<sup>3</sup>/day (B) 0.13 m<sup>3</sup>/day (C) 0.19 m<sup>3</sup>/day.

Figure 61 further illustrates the predicted effect of recirculation rate on the system by showing acetic acid concentration. With an increase of the recirculation rate, the acetic acid concentration

decreases in the high solids digester. As the acetic acid concentration falls in the high solids digester, it rises in the UASB seed reactors.

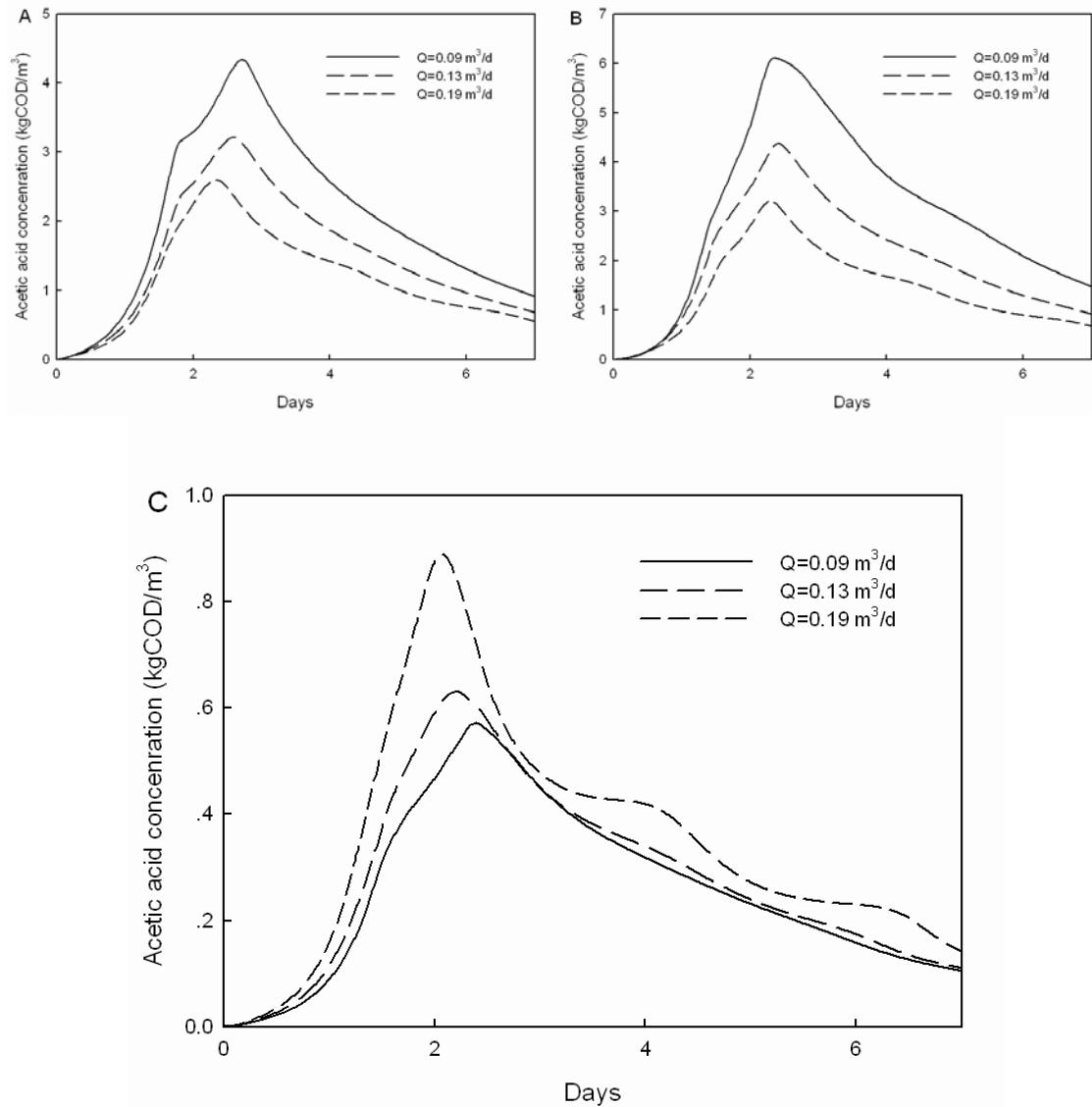


Figure 61. Predicted effect of recirculation rate on acetic acid concentration (A) HSAD (B) HSAD bottom (C) UASB.

In summary, results show that the optimized effluent recirculation rate could be predicted by the model. The balances among the volume ratio of the solids digester and the liquid digester, the biomass wash out velocity in the liquid digester, and the pH maintenance capacity by methanogens could also be determined to lower the operational cost and enhance bio-methane productivity.

## **Objective 5: Nutrient recovery after floor-scale testing of moderate solids system**

The IMARNRS technology, previously developed for use with dairy manure, was successfully used to treat effluent from the moderate solids digester system. As a first step, prior to testing effluent from the floor-scale system, the results of ammonia stripping at different TAN concentrations are shown in Table 14 (Jiang, 2010). Statistical analysis was used to show that there is no significant difference between any of the TAN concentration levels at alpha=0.05 ( $p = 0.3593$ ). This suggests that pH only affects the free ammonia ratio, and indicates that ammonia stripping efficiencies can be analyzed without consideration for changes in initial TAN concentration. Thus, conclusions from these pilot-scale tests (as well as previous bench scale tests) are quite applicable to other nutrient recovery projects.

Table 14. Ammonia stripping efficiency at different TAN concentrations.

Replicate	pH	Temperature	TAN in, mg/L	Ammonia removal
1	9.99	34.8	488	47.50%
2	9.99	34.6	480	46.70%
3	10.01	34.5	464	44.80%
1	10.02	33.8	1163	48.20%
2	10.02	34.5	1079	46.20%
3	10.03	34.2	1147	47.30%
1	10.02	34.3	1505	47.50%
2	10.02	34.5	1491	48.40%
3	10.01	34.1	1470	47.90%
1	10.01	34.2	817	46.80%
2	10.01	34.2	836	50.60%
3	10.01	33.9	811	48.30%

Table 15 shows N and P of nutrient recovery for food waste from effluent recovered from UASB buffer tank 2 after two weeks of recycling. High efficiency of the TAN removal was achieved at 93% in one day. However, the phosphate removal needed to be controlled at 6 hours. After that, a portion of the phosphate went back to effluent in different associated form, as indicated by the variable P removal rates after 6 hours. Since UASB requires low nutrients and chemicals (Zaher et al., 2009) and food waste contains significant amounts of nitrogen and phosphorus, recovering N and P eliminates the threat of inhibition in addition to generating value-added byproducts. Using the experimental data generated by this experiment, it was estimated that integrating a nutrient removal and recovery system into the effluent recycle loop of a facility processing 100,000 MT of food waste per year could produce 376 MT of nitrogen and 40 MT of phosphorus per year. This data also suggests that water can be recycled back to the high solid digester without causing ammonia inhibition, so that the system will operate stably over time without crashing.

Table 15. N and P of nutrient recovery for food waste from UASB effluent.

Aeration time (h)	P remaining (mg PO <sub>4</sub> /L)	P removal (%)	TAN remaining (mg NH <sub>3</sub> -N/L)	TAN removal (%)	Alkalinity (mg CaCO <sub>3</sub> /L)
0	0.22	0	816	0	
2	0.16	27	774	5	4413
4	0.14	36	686	16	4065
6	0.12	45	603	26	3761
8	0.13	41	461	43	3442
10	0.18	18	418	49	3168
14	0.19	14	327	60	3076
16	0.18	18	281	66	2901
18	0.13	41	208	75	2611
20	0.12	45	184	77	2655
22	0.13	41	95	88	2401
24	0.18	18	54	93	2270

## Objective 6: Disseminate scientific publications developed as a result of this project

The project results have been disseminated through scientific publications and conference presentations in addition to the project deliverables:

- Shulin Chen, Liang Yu, Jingwei Ma, Craig Frear. (2011) Towards Developing Anaerobic Digestion Based Biorefinery – Research On Food Wastes As Feedstock, AIChE Annual Meeting: Minneapolis, MN (Oral presentation).
- Liang Yu, Usama Zaher, Quanbao Zhao, Jingwei Ma, Shulin Chen. (2011) Experimental and Modeling Study of a Two-Stage Pilot Scale High Solid Anaerobic Digester System, AIChE Annual Meeting: Minneapolis, MN (Oral presentation).
- Jingwei Ma, Liang Yu, Yubin Zheng, Shulin Chen. (2011) Enhanced Food Waste Hydrolysis for Integrated Biofuel Production, AIChE Annual Meeting: Minneapolis, MN (Oral presentation).
- Liang Yu, Jingwei Ma, Shulin Chen. Numerical Simulation of Mechanical Mixing in High Solid Anaerobic Digester. BIORESOUCE TECHNOLOGY 2011, 102(2): 1012-8. See Appendix B.
- Zhanyou Chi, Yubin Zheng, Jingwei Ma, Shulin Chen. Lipid Production with Dark Fermentation Hydrogen Production Effluent Using Oleaginous Yeast *Cryptococcus curvatus*. International Journal of Hydrogen Energy, (accepted). See Appendix C.
- Liang Yu, Jingwei Ma, Craig Frear, Quanbao Zhao, Baisuo Zhao, Shulin Chen. Multiphase modeling of mixing effectiveness in anaerobic sequencing batch reactor (ASBR). Water Research, (submitted). See Appendix D.
- Jingwei Ma, Zhi-wu Wang, Shulin Chen. Evaluating rate-limiting step in anaerobic digestion of flushed dairy manure and the effect of microbial community ratio on its kinetics. (Revision). See Appendix E.
- Quanbao Zhao, Liang, Yu, Craig Frear, Shulin Chen. Optimize butyric acid and bio-hydrogen production from food waste using untreated UASB effluent as seed. (Revision). See Appendix F.

## **Section 5: Conclusions and recommendations for pilot scale design**

A floor-scale moderate solids reactor system, approximately 10 times larger than the previously tested laboratory-scale system, was built and evaluated. This system includes two main reactors: a high solids digester and a UASB reactor. The floor-scale system used an UASB because it is a high rate digester that consumes VFA and provides effluent to the high solids digester for mixing, pH control, and seeding of anaerobic microorganisms. The UASB digester also enhances overall methane productivity by allowing for higher up-flow velocities in the range of 4 - 10 meters per hour while retaining enough methanogenic bacteria to avoid washout.

A mixing strategy and a liquid-solid separation technology that have not previously been used in a moderate solids system such as ours were proposed to enhance the reactor performance and reduce the total system costs. Compared with the existing HSAD systems, the new HSAD system shows great potential in system characteristics and productivity. These advantages support future efforts to scale up this technology for commercialization.

The high solids digester in the new HSAD system was operated to maintain pH at the range of 6.0 - 8.0, the optimum pH range for methanogenic bacteria. Our experiments also confirmed that the highest VFA productivity and TS reduction is obtained at this pH range while VFA can be directly converted into methane in the next stage. Under batch mode operation, SRTs were 7 days between batches. Under continuous mode operation, SRTs were 30 days at a solids loading rate of 5.0 - 15.0 kg VS/m<sup>3</sup> day.

This system successfully treated moderate (10% - 20%) solids food waste for six months with stable operational status. The HRTs in the high solids digester were in the range of 6 to 24 hours and the organic volumetric loading rate for VFA was 15000 - 24000 mg/L at 35 °C. The floor scale experiments showed VFA removal in the first day was 38.3% and the total VFA removal was 63.4%. The COD removal in the first day was 32.0% and the total COD removal was 56.1%. These results indicate highly efficient operation of the UASB reactor, in that most of the COD and VFA removed in the first day. Although not directly tested, the literature also suggests the new HSAD system could be applied at lower psychrophilic temperatures (13 °C to 20 °C) due to the good low temperature performance of UASB. Low temperature conditions would allow this technology to be applied in long term operation in more northern latitudes with less heating, potentially providing great advantages over current systems that need to be maintained at higher temperatures.

The mixing portion of this study suggested that VFA productivity was enhanced by vigorous mixing. Considering the liquid-solid separation and the power consumption of agitators, an intermittent mixing with enough intensity to break up the biowaste clumps is recommended in the high solids digester during future scale-up work. In the UASB seed reactor, homogeneous and mild mixing enhanced mass transfer to accelerate the release of biogas, while short hydraulic retention times accelerated the biogas production rate.

A process model appropriate for the new system was developed based on the widely used ADM1 model. The high solids digester was assumed to be a combination of continuous stirred-tank reactor (CSTR) and dispersed plug flow reactor (PFR). The liquid digester was assumed to be a

dispersed PFR where UASB, a high rate digester, was used to retain more methanogens and maintain pH at the range of 6.0-8.0 in the whole system. The model parameters were adjusted using data from the floor scale experiments. The model was then used to examine the impact of changing the recirculation rate on process stability and kinetics, since this information is difficult to obtain from experiments. Results showed that the recirculation rate of 0.19 m<sup>3</sup>/day was optimal, maintaining pH at the range of 7.0-8.0 in the high solids digester. The pH in the UASB seed reactor was maintained at 7.5-8.0 and most of VFA was consumed in this reactor.

To address the concern that high amounts of N and P could inhibit the whole system, and to obtain high value-added byproducts, nutrient recovery was carried out in this project, using the IMARNRS process previously developed by WSU. High efficiency of TAN removal (93%) was achieved in one day. Phosphate removal peaked at 6 hours, after which the reaction needed to be controlled, to prevent it from re-entering the effluent in an associated form. Using experimental data, it was estimated that integrating a nutrient removal and recovery system into the effluent recycle loop of a facility processing 100,000 MT of food waste per year could produce 376 MT of nitrogen and 40 MT of phosphorus per year.

In sum, the new HSAD system provides benefits over current commercial systems used to process moderate to high solids waste, generating additional methane, high efficiency and reduced reactor volume, and high process stability. The results of the floor scale reactor system confirmed that it is feasible to continue the scale- up of this system for pilot and commercial application.

### **Additional recommendations for pilot-scale design**

Based on the investigation of the floor-scale moderate solids reactor system, a proposed flow chart for a pilot-scale moderate solids reactor system is presented in Figure 62. The pilot-scale system includes some improvements intended to further reduce system costs and enhance reactor performance. While some of these were tested in the pilot scale, others have not yet been tested, though they are suggested by the results obtained during the floor scale experimentation.

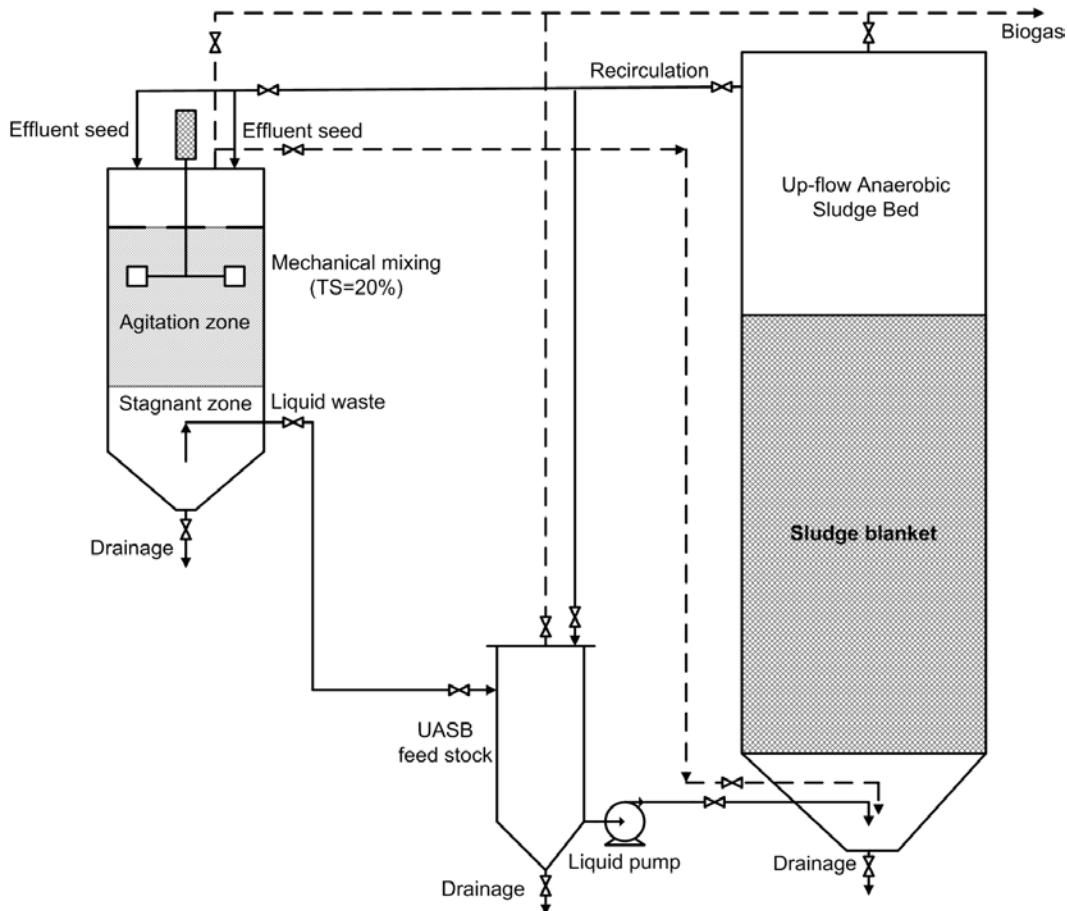


Figure 62. Flow chart of pilot scale moderate solids anaerobic digester system.

### Liquid-solid separation

When the VFA leaves the high solids digester, the VFA-containing liquid should be separated from the solid waste. Since the UASB seed reactor requires <1% TS, liquid-solid separation is critical to stabilize the new HSAD. Based on the phenomenon observed in the experiments (Figure 47), natural separation arising from the low density of food waste and the flotation of biogas should be used, as was done in the floor scale set-up. This will increase the efficiency and reduce costs compared to methods that rely on solid recycling or other additional equipment.

### Mixing strategy

Following the strategy implemented in the floor scale system (and adopting methods used in other types of AD), three different mixing modes should be used in the high solids digester to reduce energy consumption and enhance mixing efficiency. First, biogas is produced inside the media and pushes the solid waste to the top. This is natural mixing. Second, intermittent mechanical mixing should be used to break up the solid waste in the top layer. Third, the effluent from UASB seed reactor should be sprayed into the high solids digester at the top inlet. Among these three mixing modes, only mechanical mixing requires additional energy consumption. For the mechanical mixing, ordinary radial flow impellers are recommended to impose shear stress to

the fluid because this type of impeller has relatively low cost and the food waste is soft enough to be easily agitated.

### **Introduce biogas from the high solids digester into the UASB during initial system setup**

Food waste contains a higher percentage of VFA and easily degradable substrates than many other feedstocks. When the solid waste initially contacts microorganisms in the high solids digester, a high volume of CO<sub>2</sub> and H<sub>2</sub> are produced and released so rapidly that methanogens cannot capture and convert them into methane even at a pH over 6.0. In order to produce more methane, we recommend that biogas that is initially produced in the high solids digester should be captured and sent to the UASB seed reactor, where it can be retained for a longer time period.

### **Use the vertical liquid outlet in the high solids digester**

In order to effectively separate liquid and solid and avoid clogging the liquid outlet, a vertical liquid outlet design in the high solids digester is recommended as shown in Figure 62. This design effectively prevented clogging in the floor scale experiments.

### **Simplify the process through the use of gravity transport**

Compared with the floor scale design (Figure 41), the pilot scale design has been simplified by eliminating four pumps and two tanks (Figure 62), with gravity flow replacing pump transportation where feasible.

## **Public benefits to Washington**

There are four potential public benefits from the new system developed in this research project, as described in previous research (Zaher et al., 2009). The first benefit is an economic one. Previous work at the laboratory scale (incorporating a scaling factor to account for anticipated changes during scale-up) has indicated that a plant processing 17,640 tons food waste per year could save 31%, or greater than \$7.5 million annually, compared to an existing AD technology (Zaher et al., 2009). Additional refinement of these estimates was not appropriate during the current floor scale work which focused instead on technical and engineering issues. However, additional economic analysis relevant for commercial scale planning will be carried out during future pilot scale work.

Other non-economic benefits could also be expected from the new system. The second potential benefit is that the developed system nearly eliminates the emission of odors or toxic gases since the treatment is completed in a completely closed system. Third, the nutrients can be concentrated in a recycled liquid stream and recovered as fertilizers, rather than released to the environment, where they may contribute to environmental problems. Fourth, the system produces biogas of > 50% methane that can be utilized as a renewable source for energy production. Experimental data generated in this study suggests that treatment of all food waste in Washington using a HSAD system would generate 157 GWh every year.

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# Appendices

## Appendix A. Floor scale moderate dual reactor system drawing

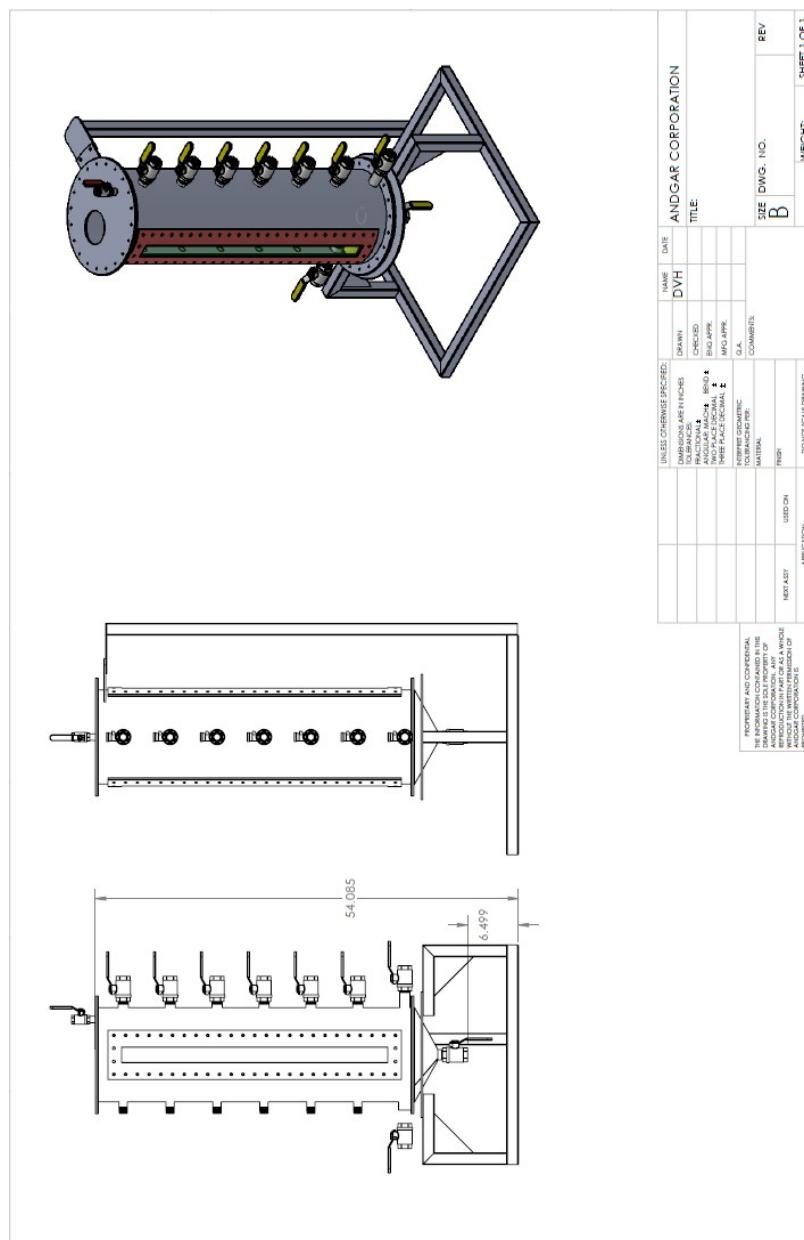


Figure A-1. High solids digester.

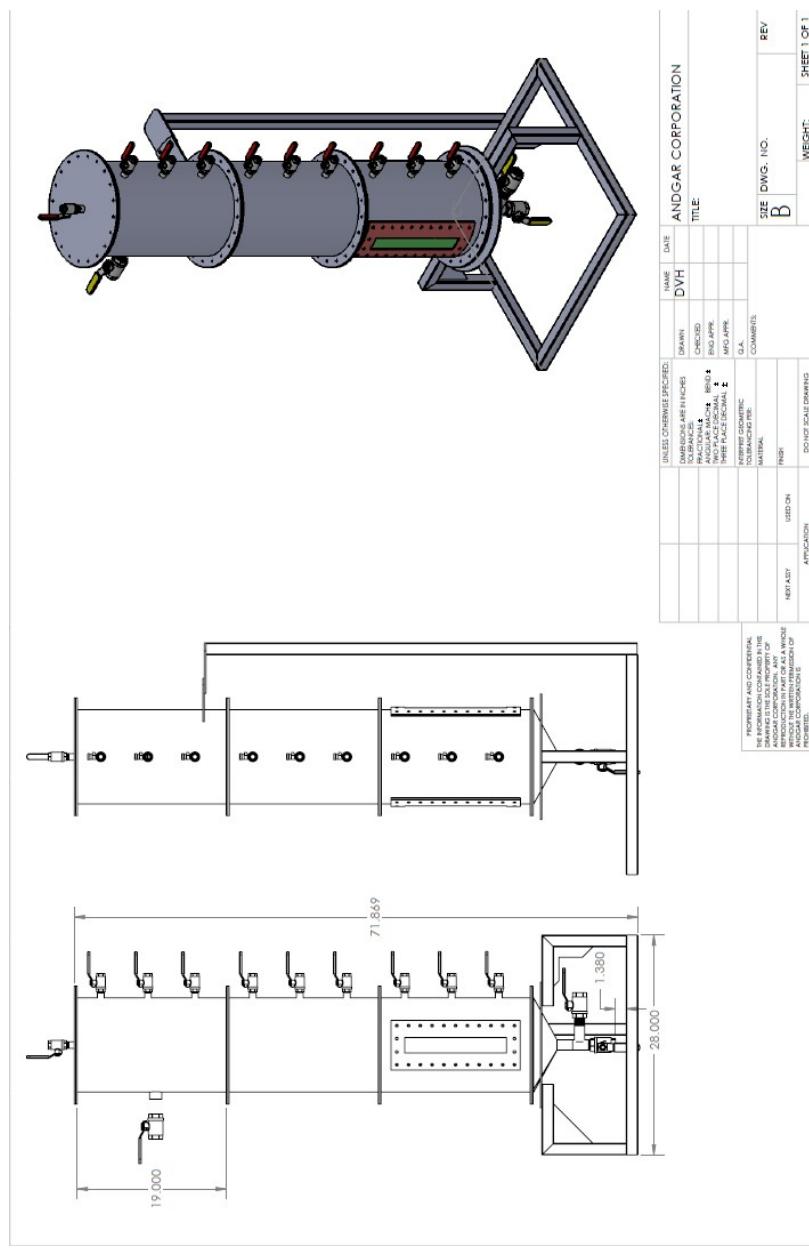


Figure A-2. Buffer tanks.

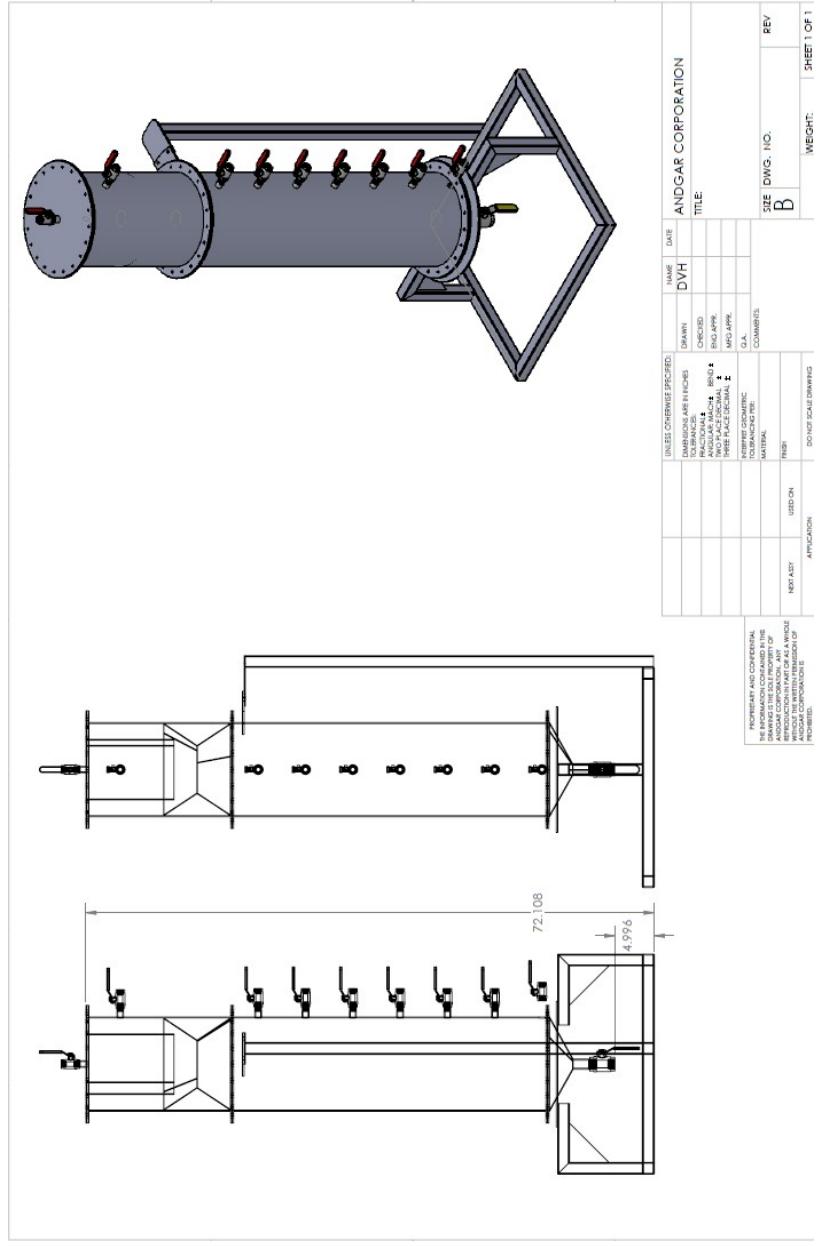


Figure A-3. Up-flow Anaerobic Sludge Beds (UASB) seed reactor.

## **Appendix B. Numerical simulation of mechanical mixing in high solid anaerobic digester**

### **Abstract**

Computational fluid dynamics (CFD) was employed to study mixing performance in high solid anaerobic digester (HSAD) with A-310 impeller and helical ribbon. A mathematical model was constructed to assess flow fields. Good agreement of the model results with experimental data was obtained for the A-310 impeller. A systematic comparison for the interrelationship of power number, flow number and Reynolds number was simulated in a digester with less than 5% TS and 10% TS (total solids). The simulation results suggested a great potential for using the helical ribbon mixer in the mixing of high solids digester. The results also provided quantitative confirmation for minimum power consumption in HSAD and the effect of share rate on bio-structure.

## **Appendix C. Lipid production with dark fermentation hydrogen production effluent using oleaginous yeast *Cryptococcus curvatus***

### **Abstract**

Volatile fatty acids (VFA) from dark fermentation hydrogen production were tested in this study as a feedstock for culture of oleaginous yeast *Cryptococcus curvatus*, which is a promising feedstock for biofuel production. First, the optimal acetate concentration and pH were investigated when potassium acetate was used as solely carbon source. Then, hydrogen production effluent (HPE) from synthetic wastewater was tested as feedstock. A pH-stat culture fed with acetic acid finally produced 168 g/L biomass, with lipid content of 75.0%. This indicates no inhibitor to yeast growth produced in hydrogen production process. However, inhibition occurred in culture with HPE from food waste (FW), indicating inhibitor may be brought from original raw material. This inhibition was avoided with a process that uses glucose as initial carbon source and continuously fed with FW-HPE. The biomass productivity in this continuous culture process reached 0.34 g/L/hour, but lipid content was only about 13.5%. These results suggest that FW-HPE is not a good feedstock, but HPE derived from nitrogen-deficient waste streams could be good feedstocks. This study provided preliminary evidence for the feasibility of lipid and hydrogen co-production from organic waste.

## **Appendix D. Multiphase modeling of mixing effectiveness in anaerobic sequencing batch reactor (ASBR)**

### **Abstract**

Effective suspension and settling are critical to increasing retention time of biomass in an anaerobic sequencing batch reactor (ASBR). In this paper, a multi-fluid model with kinetic theory of granular flow (KTGF) was developed to describe these phenomena in the pseudo-plastic biowaste particles flow in a manure-fed ASBR. A spouted bed was introduced as a mixing mechanism in ASBR. The model was experimentally validated at gas and solid volume fraction and granular temperature in a liquid-gas-solid slurry bed. The effect of restitution coefficient on rheological behavior of biowaste particles was predicted. The results showed that granular temperature decreases with an increase of the restitution coefficient and that biowaste particles tend to have fluid properties. Settling and suspension processes of ASBR were simulated for lab and pilot-scale reactors with comparisons made for reactor configuration, mixing mode, and geometry model, respectively. This study demonstrates that the multi-fluid model with KTGF could provide visually better understanding suspension and settling in ASBR. The spouted bed shows potential to be a simple internal structure within ASBR for avoidance of clogging of manure waste containing supporting media such as fiber, granule and polymers etc. The model of this study can be extended to optimize the design for other anaerobic digesters.

## **Appendix E. Evaluating rate-limiting step in anaerobic digestion of flushed dairy manure and the effect of microbial community ratio on its kinetics**

### **Abstract**

Anaerobic digestion of insoluble substrate is a complex reaction, which has been described as a multi-step process. The whole anaerobic digestion process is controlled by the rate-limiting step. A kinetics study for anaerobic digestion of flushed dairy manure was performed to evaluate the rate-liming step by addition of intermediate from each step. The experiment elucidated that hydrolysis was the rate-limiting step when anaerobic sludge was used as inoculum. Balanced anaerobic microbial community structure plays a pivotal role in an efficient anaerobic digestion process. The effects of microbial community ratio on the anaerobic degradation of dairy manure were also investigated in this study by manipulating two different inoculum, i.e. normal anaerobic sludge (NS) and heated anaerobic sludge (HS). Results revealed that the rate-limiting step changed according to the variation of microbial community ratio. It showed that there existed a critical ratio  $r^*$  of hydrolytic bacteria (H) to methanogens (M) at 24; as  $r$  decreased or exceeded from this value, hydrolysis or methanogenesis could be rate-limiting step, respectively.

## **Appendix F. Optimize butyric acid and bio-hydrogen production from food waste using untreated UASB effluent as seed**

### **Abstract**

The individual and mutual effects of total solid content, reaction time and mixing intensity on the hydrogen production by UASB effluent were investigated in a batch system. Response surface methodology (RSM) with a Box–Behnken design was employed to determine the optimum condition for enhanced hydrogen and VFA production. The hydrogen production rate was investigated by simultaneously changing the three independent variables, which all had significant influences on the hydrogen production rate. The maximum butyric acid production of 1614 mg/l was obtained under the optimum condition with TS 3.8% for 2.4 days fermentation at a mixing rate of 42 rpm. The experimental results showed that the RSM with the Box–Behnken design was a useful tool for achieving high rate of butyric acid production from food waste.