The Technical and Economic Feasibility of Producing Methane from Biomass Using a Leaching-Bed/Packed-Bed Conversion Process

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THE TECHNICAL AND ECONOMIC FEASIBILITY
OF PRODUCING METHANE FROM BIOMASS USING
A LEACHING-BED/PACKED-BED CONVERSION PROCESS

BY

STEVEN WAYNE HINTON
B. S., University of Florida, 1974

THESIS

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ABSTRACT

The economic feasibility and energy effectiveness of producing pipeline quality methane gas from biomass was assessed for a new and totally unproven process. The biomass feedstock considered was the common aquatic weed water hyacinth and a novel active boom-winches harvesting system is proposed for its collection. The conversion process analyzed is a two stage biological process which utilizes a leaching-bed for the production of volatile acids and a packed-bed for the production of methane gas. In order to determine the feasibility of the proposed process equipment requirements, capital costs and operating/maintenance costs were developed for three system sizes. This data was analyzed using a life cycle cost model to determine payback period. The results indicate that payback period will be less than equipment life and that net energy production occurs. Areas where further research would promote the introduction of this technology are identified and discussed.
ACKNOWLEDGEMENTS

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LIST OF SYMBOLS

A = area (ft$^2$)

An = normal area (ft$^2$)

ARC = annual recurring cost ($/yr)

At = tangential area (ft$^2$)

$C_1$ = conversion constant ($144 \text{ in}^2/\text{ft}^2$)

$C_2$ = conversion constant ($0.01078 \text{ ft}^2-\text{sec}^2/\text{in}^2$)

$C_3$ = conversion constant (192.5 ft$^3$-min/gal-day)

$C_4$ = conversion constant (11,936 lb-min/gal-day)

$C_5$ = conversion constant (3,960 gal-ft/hp-min)

$C_6$ = conversion constant (1.8 ton-sec/hr-lb)

$C_7$ = conversion constant (0.001 hp-hr/ton-ft)

$C_8$ = conversion constant (2,000 lb/ton)

$C_9$ = conversion constant (43,560 ft$^2$-day/acre)

$C_{10}$ = conversion constant (550 ft-lb/sec-hp)

$C_{11}$ = conversion constant (365 days/yr)

$C_{12}$ = conversion constant (3,600 sec/hr)

$C_{13}$ = conversion constant (2,545 BTU/hp-hr)

$Cd$ = drag coefficient (dimensionless)

$CB$ = belt flexing power constant (hp-min/ft$^2$)

$CF$ = compaction factor for leaching-bed (dimensionless)
Const. = construction cost ($)
CR = conversion rate (SCF/lb volatile solids)
D = diameter (ft)
DATE = last period in calculation (yr)
DF\textsubscript{1} = form drag force due to wind (lb)
DF\textsubscript{2} = friction drag force due to wind (lb)
DF\textsubscript{3} = form drag force due to water (lb)
DF\textsubscript{4} = friction drag force due to water (lb)
DFT = total drag force (lb)
DP = pressure drop (lb/in\textsuperscript{2})
DP' = pressure drop per 100 feet of pipe (lb/in\textsuperscript{2})
DID = daily feedstock requirement (dry ton/day)
DIY = yearly feedstock requirement (dry ton/yr)
E = yearly electrical consumption (BTU/yr)
E\textsubscript{1} = escalation rate 1981 to 1985
E\textsubscript{2} = escalation rate 1985 to 1990
E\textsubscript{3} = escalation rate 1990 to infinity
Ec = yearly energy consumption of conveyors (hp-hr/yr)
Eg = yearly energy consumption of grinder (hp-hr/yr)
EL = effective lift (ft)
Ep = yearly energy consumption of pumps (hp-hr/yr)
Ew = yearly energy consumption of winches (hp-hr/yr)
Ex = yearly energy consumption of compressor (hp-hr/yr)
f = fanning friction factor (dimensionless)
fd = discharge friction head (ft)
fs = suction friction head (ft)
FDM = fraction dry matter (dimensionless)
FFC = future fuel cost ($/10E6 BTU)
FFCPW = future fuel cost present worth ($/10E6 BTU)
FSPWS = fuel series present worth sum ($-yr/10E6 BTU)
FVS = fraction volatile solids (dimensionless)
g = acceleration of gravity (32.2 ft/sec²)
GPM = flow rate (gal/min)
h = enthalpy (ft-lb/lb)
hc = BTU content of gas (1,000 BTU/ft³)
hd = static discharge head (ft)
hs = static suction head (ft)
H = total head (ft)
HA = harvest area (acre/day)
i = annual interest rate ($/$)
K = ratio of specific heats (dimensionless)
l = length (ft)
L = effective pipe length (ft)
LBD = weight flow rate (lb/day)
m = mass flow rate (lb mole/sec)
n = number of years (yr)
Ne = electric motor efficiency (dimensionless)
Np = pump efficiency (dimensionless)
Nw = winch efficiency (dimensionless)
Nx = compressor efficiency (dimensionless)
p = weight density (lb/ft³)
ps = static pressure, absolute (lb/ft²)
pw = wet weight density (lb/ft³)
P = power (ft-lb/sec)
PAT = percent active time of leaching-bed (dimensionless)
Pc = input power to conveyor (hp)
Pe = shaft power @ conveyor to move empty belt (hp)
Ph = shaft power @ conveyor to move load horizontally (hp)
Pp = input power to pumps (hp)
Pt = total shaft power @ conveyor (hp)
Pv = shaft power @ conveyor to raise load vertically (hp)
Pw = input power to winches (hp)
Px = input power to compressor (hp)
Q = heat added to fluid (ft-lb/lb)
R = ideal gas constant (1,544 ft-lb/lb mole-R)
S = belt speed (ft/min)
SCF = gas volume @ standard conditions (ft³)
SD = standing plant density (wet ton/acre)
SPFW = single payment future worth ($/$)
SPPW = single payment present worth ($/$)
t = time (hr/yr)
T = temperature (R)
TC = towing constant for triangle fann (ft/day-ft)
TFC = today's fuel cost ($/10E6 BTU)
TH = belt conveyor capacity (ton/hr)
u = internal energy of fluid (ft-lb/lb)
USPW = uniform series present worth ($-yr/$)
v = specific volume (ft³/lb)
V = velocity (ft/sec)
VD = volume of mass digested (ft³/yr)
VV = vessel volume of leaching-beds (ft³/yr)
w = weight flow rate (lb/sec)
wd = belt width (in)
W = work done on fluid (ft-lb/lb)
W' = work done on fluid (ft-lb/lb mole)
WG = grinding work constant (hp-hr/green ton)
WHP = work horsepower (hp)
x = methane flow (SCF/day)
X = plant BTU output per year (BTU/yr)
XS = load's cross sectional area (ft²)
y = volatile acid flow (lb/day)
Y₁ = number of years in period 1981 to 1985 (yr)
Y₂ = number of years in period 1985 to 1990 (yr)
Y₃ = number of years in period 1990 to infinity (yr)
YEAR = calculation period (yr)
Z = elevation above datum (ft)
I. INTRODUCTION

In recent years consumers of fossil fuels have encountered supply interruptions and large price increases. This combined with a growing awareness that the world's fossil fuel resources are finite has prompted considerable research into renewable energy sources. Much of this research has focused on developing fundamental data without detailed consideration of how an integrated system might function. As the data base becomes more defined, new and unproven methane production concepts are beginning to emerge. These concepts are analyzed for economic viability and energy effectiveness to identify key problem areas where future research efforts could provide the largest benefit. Focusing research on these key areas should promote fast introduction of the new technologies as well as lower the cost of these new energy sources.

The expected United States energy consumption for 1983 is $78 \times 10^{15}$ BTU per year [1]. Of this amount 27.5 percent or $21.5 \times 10^{15}$ BTU per year will be natural gas (methane) consumption. At this time the United States is nearly self-sufficient in natural gas as a whole but most of the gas production is geographically located in just a few regions of the country such as the Gulf coast, the Rocky Mountain area and the Appalachian area. The gas is drawn
from the ground and pumped in pipelines to areas like Florida which have essentially no known gas reserves. In the early 1970's the United States gas consumption outstripped new gas discoveries and the total U.S. reserve of gas began to decline. In the late 1970's however, legislation was passed which provided economic incentives for more advanced types of gas exploration. The result was that new gas discoveries exceeded consumption and the net proven reserves started to increase. This positive benefit came with a cost. The average wholesale price of natural gas to a typical Florida gas utility [2] increased 2.6 times between October 1978 and April 1982.

Although the near term supplies of natural gas appear assured, the long term outlook is one of dwindling reserves. This is true not only for United States gas supplies but for all the world's fossil fuel resources. Simply put, all fossil fuels were formed millions of years ago by a geological and biological action. The result of these actions was the formation of a finite resource. Once expended this resource will be gone forever. Unconventional exploration and recovery methods may help maximize the utilization of this finite supply but ultimately the entire resource will be expended.

The utilization of natural gas is a diverse activity which has penetrated all levels of the United States economy. Gas is used for the everyday heating of homes and businesses as well as powering factories and generating electric power. This was made possible by the construction of an elaborate system of underground transmission pipelines and distribution networks. These systems along with the
equipment which consumes gas represents a sizable investment in the United States economy. As the existing supply of natural gas becomes unable to meet the demand, the most convenient way to fill this energy deficit would be to provide substitute natural gas (SNG) to the existing energy delivery/consumption system. This method of gradually integrating SNG into the market place should produce the least amount of shock to the economy as a whole. Ideally the SNG should come from a renewable source like biomass. Here biomass is defined as any plant material which can be converted to natural gas by either thermal or biological means. Typical examples of biomass would include wood, herbaceous plants, macro algae, etc. Biomass sources would include cultivated energy crops as well as waste materials.

Methane can be produced either by thermal processes or biological processes. Thermal gasification involves the heating of biomass in the absence of oxygen. This results in a gas product which is approximately 50 percent methane [3]. Biological gasification involves subjecting the biomass to bacteria which produce methane gas as a by-product of their existence. This can result in a gas product which is 85 percent methane [4]. By comparison, the composition of present day pipeline gas is approximately 97 percent methane. This study will examine possible alternatives to producing pipeline quality methane from biological processes.

The formation of methane through a biological process is commonly called anaerobic digestion. It is a microbial action in the absence
of oxygen and the overall chemical reaction is represented by

$$\text{organic matter} - \overset{\text{anaerobes}}{\rightarrow} \text{CH}_4 + \text{CO}_2 + \text{H}_2 + \text{N}_2 + \text{H}_2$$

(1)

The digestion process is actually a complex chain of reactions involving a number of reaction steps and alternate pathways [5]. Researchers have found that separate microbial populations are catalysts for a 3-step reaction process. First, a group of acid-forming bacteria solubilizes the volatile solids (the part which can be converted to gas) and then converts them to volatile organic acids. Finally, a group of methanogenic bacteria converts the volatile acids to methane and carbon dioxide. Hence Equation 1 can be recast to show the 3-step reaction as follows:

insoluble organic matter $\overset{\text{acid formers}}{\rightarrow}$ soluble organic matter (2)

soluble organic matter $\overset{\text{acid formers}}{\rightarrow}$ volatile acids (3)

volatile acids $\overset{\text{acid formers}}{\rightarrow}$ \text{CH}_4 + \text{CO}_2 + \text{miscellaneous}$ (4)

The proven methods for the production of methane evolved from sewage disposal technology. They utilize a process which combines both the acid formers and the methane formers in a single 'stage' reactor. A single-stage reactor is a sealed vessel which contains a mixed population of methanogenic and acid forming anaerobes, combined with the biomass to be digested. There the three steps outlined in Equations 2, 3, and 4 occur simultaneously and in association with each other. Typical efficiencies for single-stage processes range from $n = .15$ to $.50$. Attempts to improve efficiency have only been
partially successful because the two chemical reactions do not optimize on the same parameters. Acid forming anaerobes prefer a pH less than six and short solids retention times. Methanogenic anaerobes on the other hand prefer a pH near seven and long solids retention times. Solids retention time is the time microbes are allowed to feed and multiply before they are removed from the reactor.

In addition to low conversion efficiencies, the proven processes have frequently been operated thermophilically (at 131°F) with mechanical mixing. The purpose of heat and mixing is to obtain reliable operation and not to increase gas production. The reason for this is that these techniques were developed to dispose of sewage and not produce gas. Heat and mixing, although they promote consistent gas production, further reduce the overall process efficiencies.

In an attempt to improve the biological as well as the overall process efficiencies, several investigators have recently proposed conversion concepts which would divide the methane production process physically into two stages. In a two phase process, the acid-forming and methanogenic anaerobes are cultured in separate vessels where conditions can be independently tailored to suit each population. Biomass is fed to the acid-phase digestor (APD) where volatile solids are solubilized and converted to volatile acids. The volatile acid stream is then fed at controlled rates to the methane phase digestor (MPD). Preliminary laboratory results indicate that conversion efficiencies for this concept will approach 85 percent. This has prompted speculation that a two-stage process could provide competitively
priced SNG. Presently however, the true potential of two-stage digestion is unknown because an integrated system analysis has not been completed. This investigation seeks to separate the real potential from the perceived potential by determining both the economic feasibility and the energy effectiveness of the most promising two-stage conversion process.
II. BIOMASS TO METHANE CONVERSION ALTERNATIVES

Since commercial biomass to methane conversion systems do not exist, much of this analysis is based on data from sewage treatment processes. Anaerobic digesters for conventional sewage treatment have been designed in a variety of configurations and operating modes including batch and continuous systems. This study considers continuous and semi-continuous systems only. The five general configurations of anaerobic digesters are as follows:

1. slurry reactors
2. fixed-film reactors
3. sludge blanket reactors
4. batch reactors
5. expanding bed reactors

Each configuration has the potential to be used as either an acid phase digester (APD) or a methane phase digester (MPD) in a two-stage system. The characteristics of each configuration will be discussed below.

1. **Slurry Reactor**

A slurry reactor is operated by mixing small particles of solid substrate (the material to be digested) with water to a pumpable consistency and feeding the slurry continuously into a closed, agitated vessel containing slurry that is digesting. Fresh feed is continuously mixed with the digester contents, and effluent is
continuously removed. This design provides close contact between substrate and anaerobes. The mixing also prevents scum formation and upsets from toxic slugs. Mixing can be accomplished by mechanical agitation, gas injection, effluent recycle, or jacketed draft tube circulation.

The major drawback to slurry systems is their high energy requirements for mixing and heating. In addition, to be efficient the slurry process requires feed material with a particle size of approximately 0.5 inch. This requires additional energy inputs for chopping and grinding as most feedstocks are at least one or two orders of magnitude larger than the desired size.

2. **Fixed-Film Reactor**

The fixed-film process depends on developing a growing population of microorganisms attached to a solid support. There are three general types of fixed-film systems as characterized by their supporting structure: (1) rotating biological contractors (Figure 1) which consist of plastic disks, mounted on a rotating horizontal shaft, (2) packed-bed reactors (Figure 1) which are vertical vessels packed with inert solids of high percentage void spaces, (3) fluidized bed reactors (Figure 1) which are vertical vessels packed with inert solids of low percentage void spaces. In all of the configurations, the material to be digested is fed continuously across the mass of microbes.
a) **ROTATING BIOLOGICAL CONTACTORS**

b) **PACKED BED REACTOR**

c) **FLUIDIZED BED REACTOR**

d) **SLUDGE BLANKET REACTOR**

Figure 1. Typical Reactor Configurations
Fixed-film reactors provide a balance between biomass conversion, energy consumption and capital costs. The plug flow regime approximated by these systems favors high overall conversion for small equipment sizes, but makes these systems susceptible to failure caused by physical plugging and toxic slugs. To minimize these shortcomings, solids concentrations within the reactor stream are kept low and the substrate is recycled through the vessel by pumping. Energy inputs for this process are moderate due to recycling of the substrate through the solid supports.

3. **Sludge Blanket Reactor**

   This is a suspended growth digestion process which relies on the formation of a dense, stable anaerobic sludge blanket in the middle section of the reactor. Substrate is fed up through the sludge blanket into an upper quiescent zone which allows entrained solids to settle back onto the blanket. See Figure 1 for configuration details.

   The key to this system is the formation of a highly active, easily settleable sludge mass. Although these systems can provide good conversion rates with low energy inputs and moderate capital costs, they appear not suitable for cellulosic feeds which have a tendency to float.

4. **Batch Reactor**

   A batch digester is a vessel loaded with biomass, sealed and allowed to ferment undisturbed. When biological activity ceases, the
digester is emptied and the process is repeated. Biological activity is not a continuous process as is shown in Figure 2. Complete conversion of the material can take up to one to six months depending on whether the reactor is used for methane production or volatile acids production. The batch reactor concept utilizes the same capital equipment as the expanding bed concept to be discussed next. Based upon capital dollar invested, the batch reactor is not as productive as other alternatives because of the low gas production rates at the end of its cycle.

5. Expanding-Bed Reactor

This configuration utilizes the same equipment as the batch digester but it has a higher output (producing either volatile acids or methane) per capital dollar invested. As biological action occurs, the volume of solids in the reactor decreases. At some predetermined (optimized) time the reactor is opened and additional feedstock material is added without clean-out of the reactor. As the solids' volumes is greatly reduced by the biological activity, clean-out is rarely expected and the output of the reactor should be nearly continuous.

The expanding bed system can be utilized for either volatile acids production or methane gas production. The first reaction occurs rapidly while the second reaction is very sluggish. For this reason expanding bed systems are most suitable for volatile acids production and when configured for this purpose they are referred to as
Figure 2. Batch Digestion - Typical Production Curve*

* Data from B. C. Wolverton and R. C. McDonald. Energy From Vascular Plant Wastewater Treatment Systems. Earth Resources Laboratory, National Aeronautics and Space Administration National Space Technology Laboratories, NSTL Station, Mississippi, 1980.
"leaching-bed digesters" (see Figure 3). Neither the technical nor economic feasibility of the leaching-bed has been demonstrated but conceptually this process has many superior advantages. Equipment requirements are simple and only an open topped vessel is required because the volatile acids are always in the liquid phase. Operationally the process is stable and requires only a minimum of supervision and controls. Finally, energy inputs are at a bare minimum as only pumping energy to raise the substrate to the bed's top is required.

Based on the discussion above, Table 1 was created to rank possible system configurations. Factors considered important in the selection of the candidate system are as follows:

1. The process should be suitable to biomass feedstock in its most natural form. For volatile acids production suitability implies a minimum of chopping before introduction into the reactor while suitability for methane production implies short reaction times.

2. Equipment complexity should be minimized so as to provide low capital and maintenance costs.

3. Controllability signifies the tendency of a process to return to design conditions after a process upset. Systems with a high degree of controllability tend to be naturally stable and therefore require only a minimum of supervisory personnel and controls.

4. Energy input should be minimized so as to provide low operating cost and a high energy output to energy input ratio.
Figure 3. Leaching-Bed Reactor
# TABLE 1

**PROCESS COMPARISON - BIOMASS TO METHANE CONVERSION ALTERNATIVES**

<table>
<thead>
<tr>
<th>Process</th>
<th>Volatile Acid Production</th>
<th>Methane Gas Production</th>
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<tr>
<td></td>
<td>Suitable For Biomass Feedstock</td>
<td>Equipment Complexity</td>
</tr>
<tr>
<td>Slurry</td>
<td>M</td>
<td>H</td>
</tr>
<tr>
<td>Fixed-film</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rotating biological contractor</td>
<td>M</td>
<td>H</td>
</tr>
<tr>
<td>Packed-bed</td>
<td>M</td>
<td>M</td>
</tr>
<tr>
<td>Fluidized bed</td>
<td>M</td>
<td>M</td>
</tr>
<tr>
<td>Sludge blanket</td>
<td>L</td>
<td>L</td>
</tr>
<tr>
<td>Batch reactor</td>
<td>--- not continual ---</td>
<td>--- not continual ---</td>
</tr>
<tr>
<td>Expanding bed</td>
<td>H</td>
<td>L</td>
</tr>
<tr>
<td><em>(leaching-bed)</em></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**SOURCE:** L. J. Bilello, Senior Engineer, Environmental Science and Engineering, Unpublished Data, Gainesville, Florida 1982.

**NOTES:** L = low, M = moderate, and H = high.
The rankings of Table 1 show that an excellent two-stage biomass to methane conversion process might be obtained by using a leaching-bed acid phase digester followed by a packed-bed methane phase digester. The leaching-bed was selected for the acid phase because it is technologically simple, stable, non-energy intensive and biologically efficient. The packed-bed was selected for the methane phase because it is the least energy intensive process which is stable and biologically efficient.
III. CANDIDATE SYSTEM ANALYSIS

This chapter outlines the assumptions and methods used to analyze a two-stage biomass to methane conversion system. Part of this analysis examines the effects of scale on gas production cost and net energy production. Scale effects are improvements due to a change in size (e.g. A 200 gallon tank costs less than the price of two 100 gallon tanks). In an effort to simplify the presentation of the study methodology, this chapter discusses the facts relevant to a 0.1(10E12) BTU per year conversion facility. Ultimately Chapter IV will present analyses of 0.1, 0.5 and 1.0(10E12) BTU per year systems. The analysis technique is the same for all three sizes and for the most part, the data and assumptions for the two larger sizes are simply multiples of the 0.1(10E12) BTU per year system. The differences that do exist are small and are discussed in Chapter IV with the results.

Previously it was shown that a two-phase conversion process which utilized a leaching-bed followed by a packed-bed represents an optimum biomass to methane system. Although this arrangement has never been tested, bench scale tests by Colleran [4] have successfully produced 6.0 standard cubic feet (SCF) per pound volatile solids applied. These tests were conducted in vessels which were exposed to ambient weather conditions somewhat cooler than Florida. Although detailed energy consumption data was not collected, the experimenter observed
that the chemical reactions occurring in the leaching-bed were sufficiently exothermic to provide all process heat losses.

**Process Description**

Table 2 and Figure 4 represent a material balance and a flow diagram respectively of a leaching-bed/packed-bed methane production system. This system was proposed by Bilello [6] for the express purpose of producing methane gas from biomass. Significant stream flows are identified on Figure 4 by boxed numbers (e.g. □). Table 2 is arranged as a matrix of rows and columns. The numbered columns correspond to the stream flows on Figure 4 and have the following designations:

1. biomass feedstock supply
2. volatile acids supply
3. lime slurry supply
4. volatile acids feed
5. MPD liquid effluent
6. MPD gas effluent
7. APD liquid effluent
8. liquid waste
9. plant solid waste effluent
10. plant liquid waste effluent
11. lime slurry make-up water
12. supply water

The rows of Table 2 detail the composition of each stream flow. Row designations are as follows:

1. total solids (lb/day)
2. volatile solids (lb/day)
3. volatile acids (lb/day)
4. water (lb/day)
5. total (lb/day)
6. methane (standard ft³/day)
7. carbon dioxide (standard ft³/day)
Figure 4. Flow Diagram - Leaching-Bed/Packed-Bed Reactor Process
### TABLE 2

**MASS BALANCE - 0.1(10E12) BTU/yr**

*(Figures in 100,000)*

<table>
<thead>
<tr>
<th>STREAM NUMBERS</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
<th>10</th>
<th>11</th>
<th>12</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Total Solids <em>(lb/day)</em></td>
<td>0.610</td>
<td>0.030</td>
<td>0.030</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.232</td>
</tr>
<tr>
<td>2. Volatile Solids <em>(lb/day)</em></td>
<td>0.519</td>
<td>0.030</td>
<td>0.030</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.140</td>
</tr>
<tr>
<td>3. Volatile Acids <em>(lb/day)</em></td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
<td>0.378</td>
</tr>
<tr>
<td>4. H₂O <em>(lb/day)</em></td>
<td>11.590</td>
<td>12.600</td>
<td>0.270</td>
<td>12.670</td>
<td>12.870</td>
<td>0.000</td>
<td>23.219</td>
<td>10.915</td>
<td>1.160</td>
<td>10.915</td>
<td>0.270</td>
<td>0.270</td>
</tr>
<tr>
<td>5. Total <em>(lb/day)</em></td>
<td>12.200</td>
<td>12.978</td>
<td>0.300</td>
<td>13.278</td>
<td>13.085</td>
<td>0.000</td>
<td>23.597</td>
<td>10.915</td>
<td>1.392</td>
<td>10.915</td>
<td>0.270</td>
<td>0.270</td>
</tr>
<tr>
<td>6. CH₄ <em>(SCF/day)</em></td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>3.11</td>
</tr>
<tr>
<td>7. CO₂ <em>(SCF/day)</em></td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
<td>0.549</td>
</tr>
</tbody>
</table>

**SOURCE:** L. J. Bilello, Senior Engineer, Environmental Science and Engineering, Unpublished Data, 1982
A complete process description can be found in the Appendix. It should be noted that the material balance assumes a 0.88 plant service factor. Service factor is the percentage of time that the facility operates at full production. Therefore, a conversion facility which has a material flow like Table 2, will produce $0.1(10^{12}) \text{ BTU}$ in 7,709 of a year's 8,760 hours. In the following discussion equipment will be sized using the flow rates of Table 2 while energy consumption will be estimated using the total yearly flow required to produce $0.1(10^{12}) \text{ BTU}$ (i.e. Table 2 times 0.88).

The material balance (Table 2), the flow diagram (Figure 4), and the observations of Colleran on heat losses form the assumptions of this analysis. These assumptions will be utilized to develop capital and operating cost budgets as well as to define net energy production. From these developments the economic feasibility and energy effectiveness of the proposed methane production system will be assessed.

The biomass feedstock assumed for this analysis is water hyacinth (Eichornia crassipes; see Figure 5) which is an extremely prolific fresh water aquatic plant. Water hyacinth occurs naturally in Florida and is an ideal candidate feedstock because it simply floats on the water surface. Harvesting amounts to just scooping the plants off the water surface and transporting them to the conversion facility. The candidate farming system assumed for this analysis was proposed by Bruderly [7] and is shown in Figure 6. It is called the active boom-winches harvester system. The system uses a flexible floating boom collector positioned by four shore-based traction winches to scoop-up
Figure 5. Water Hyacinth Plant
Figure 6. Active Boom Winch Harvesting System
a day's supply of feedstock at a time. The feedstock is forced to
shore where a floating harvester lifts the material off the water's
surface, chops it into fine pieces and places the water hyacinths on a
belt conveyor. The conveyor carries the ground material to a pumping
station where it is delivered to the conversion facility via a
pipeline.

Design Considerations

The design criteria for the candidate systems involve the six
generic areas. They are as follows:

1. pipes
2. vessels
3. pumps
4. conveyors
5. winches
6. gas cleaning and compressor

The following discussion summarizes the assumptions and considerations
which established equipment sizing standards for each of the six gen-
eric areas.

1. Pipes

Pipes are the conveyance in which liquids and gases are moved
from one location to another. The motive force associated with this
action is pressure difference. Piping was sized to minimize pressure
losses without undue expenses of capital. This objective was selected
to promote high overall conversion efficiencies and can be achieved by
specifying an average frictional pressure loss of less than 1.0 psi
Pressure loss in a pipe is predicted by the familiar Darcy formula or

\[ DP = \frac{p f V^2 L}{D} \frac{2g}{C_1} \]  \hspace{2cm} (5)

where

- \( DP \) = pressure drop (lb/in²)
- \( g \) = acceleration of gravity (32.2 ft/sec²)
- \( f \) = friction factor (dimensionless)
- \( p \) = weight density of fluid (lb/ft³)
- \( D \) = pipe diameter (ft)
- \( V \) = average fluid velocity (ft/sec)
- \( L \) = effective pipe length (100 ft)
- \( C_1 \) = constant (144 in²/ft²)

By assuming all fluids are at 86°F temperature and all pipes have frictional characteristics similar to that of steel pipes, Equation 5 can be reduced to

\[ DP' = \frac{p f V^2}{D} \times C_2 \]  \hspace{2cm} (6)

where

- \( DP' \) = pressure drop per 100 ft pipe (lb/in²)
- \( C_2 \) = constant (.01078 ft²·sec²/in²)

Setting \( DP = 1.0 \) psi, setting \( D = .33 \) feet (assumed average pipe size) and solving Equation 6 for velocity provides the following allowable velocities:

<table>
<thead>
<tr>
<th>Material</th>
<th>Assumed ( f )</th>
<th>Velocity (ft/sec)</th>
</tr>
</thead>
<tbody>
<tr>
<td>( H_2O ) liquid</td>
<td>.016</td>
<td>5</td>
</tr>
<tr>
<td>water hyacinth slurry</td>
<td>.300</td>
<td>5.0 [9]</td>
</tr>
<tr>
<td>( CO_2 ) @ 1 ATM</td>
<td>.016</td>
<td>130</td>
</tr>
<tr>
<td>( CH_4 ) @ 1 ATM</td>
<td>.016</td>
<td>215</td>
</tr>
<tr>
<td>( CO_2 ) @ 17 ATM</td>
<td>.016</td>
<td>35</td>
</tr>
<tr>
<td>( CH_4 ) @ 17 ATM</td>
<td>.016</td>
<td>55</td>
</tr>
</tbody>
</table>
It should be noted that the recommended velocity for the water hyacinth slurry is a minimum of 5 feet per second and was not calculated from Equation 5. Although no data exists for this fluid, paper stock slurries require a minimum of 5 feet per second to prevent fall-out of solids [9]. Water hyacinth slurry, since it is a cellulose based material is assumed to have similar fluid properties to paper stock slurry.

2. Vessels

Vessels provide the location where the biological reactions take place. Their size is a function of reaction time and desired processing rate of feedstock material. To produce a given product (e.g. $x$ SCF methane per day) the feedstock material (e.g. $y$ lb volatile acids per day) must reside in the vessel for a time which is sufficient for the reaction to occur. Hydraulic retention time (HRT) is the ratio of vessel volume to volumetric flow rate and is the average length of time that the liquid feedstock is present in the digester. It should be long enough to allow a reasonable degree of conversion but not so long as to make the reactor size uneconomical. For all vessels except the leaching-bed reactors, this study utilizes a HRT of 1.5 days as suggested by Andrews [10]. The 1.5 days HRT recommendation is based primarily on experiences with sewage disposal systems.

The leaching-beds require a somewhat different sizing strategy because they are a semi-batch process. Feedstock is added to the leaching-beds periodically and this results in a residual solids build-up in the reactor. The residual solids eventually reduce the
active volume of the reactor to a point that the loss of production from a clean-out shutdown is less than the loss of production from reduced active volume. The required volume for leaching-beds is equal to the product of mass volume to be digested times percent active time of the reactor divided by the compaction factor for solids volume reduction. The mass volume to be digested is obtained by applying the unit process conversion factors of

1. fraction volatile solids of dry feedstock matter
2. conversion rate of volatile solids to gas
3. fraction dry matter of total wet weight
4. density as received (wet)
5. BTU content of gas

to the design plant output. The conversion factors for water hyacinth are shown in Table 3. The percent active time is the fraction of the year that volatile acids are actually produced in the reactor. For this analysis, Bilello [6] suggests a 60-day operating cycle as follows:

1. 5-day loading period
2. 30-day reaction period
3. 20-day residual solids drying period
4. 5-day loading period

The above recommendation is based on the studies of Hashimoto [11] and Colleran [4] and provides a percent active time of 0.167.

The compaction factor is the number of times the reactor can be loaded before the residual solids clean-out. Since there is no data on the compaction factor available, a conservative value of 1.5 [6] will be used in this analysis. The value is conservative because it tends to oversize vessels and therefore overstate the system's capital cost.
### TABLE 3
WATER HYACINTH PROPERTIES AND CONVERSION FACTORS

<table>
<thead>
<tr>
<th>Property</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fann yield (1)</td>
<td>33.3 dry ton/acre-yr</td>
</tr>
<tr>
<td>Standing density (1)</td>
<td>166.5 wet ton/acre</td>
</tr>
<tr>
<td>Fraction dry matter (2)</td>
<td>0.05</td>
</tr>
<tr>
<td>Unchopped density (3)</td>
<td>15 wet lb/ft³</td>
</tr>
<tr>
<td>Chopped density (4)</td>
<td>50 wet lb/ft³</td>
</tr>
<tr>
<td>Fraction volatile solids (2)</td>
<td>0.86</td>
</tr>
<tr>
<td>Conversion rate (5)</td>
<td>6.0 SCF methane/lb volatile solids</td>
</tr>
<tr>
<td>Height above water (3)</td>
<td>2.5 ft</td>
</tr>
<tr>
<td>Root length (3)</td>
<td>1.5 ft</td>
</tr>
</tbody>
</table>

**SOURCES:** Data obtained from the following:

3. P. A. Smith, Mechanical Harvesting of Aquatic Plants, Report 2, Environmental Laboratory, U. S. Army Engineer Waterways Experiment Station, Vicksburg, Mississippi, 1980.
In summary the required net leaching-bed volume per year is

\[ V_V = \frac{V_D \times PAT}{CF} \]  

(7)

where

\[ V_D = \frac{X}{FVS \times CR \times FDM \times pw \times hc} \]  

(8)

\[ V_V = \text{vessel volume of leaching beds (ft}^3/\text{yr)} \]
\[ V_D = \text{volume of mass digested (ft}^3/\text{yr)} \]
\[ X = \text{plant BTU output (BTU/yr)} \]
\[ FVS = \text{fraction volatile solids (dimensionless)} \]
\[ CR = \text{conversion rate (SCF/lb volatile solids)} \]
\[ FDM = \text{fraction dry matter (dimensionless)} \]
\[ pw = \text{wet density (lb/ft}^3) \]
\[ hc = \text{BTU content of gas (1,000 BTU/ft}^3) \]
\[ PAT = \text{percent active time (dimensionless)} \]
\[ CF = \text{compaction factor (dimensionless)} \]

Since the gross volume of the leaching-beds must also contain the gravel bed sump and the recycle spray system (Figure 3), the design volume of the leaching-beds must be somewhat larger than Equation 7 predicts. This analysis will allow 10 feet of the vessel's height for this nonreactive volume [6]. This allowance is typical for similar equipment utilized in conventional sewage treatment. The BTU content of the produced gas can vary from about 600 BTU per standard cubic foot to 1,050 BTU per standard cubic foot depending on the amounts of non-methane constituents left in the gas after processing. A value of 1,000 BTU per standard cubic foot was assigned because it approximates pipeline quality standards and it simplifies future calculations. Details of the gas cleaning process will be discussed shortly.

3. Pumps

Pumps are the motive force which move the feedstock material through the conversion process. Their size is a function of the pump
head times weight flow rate product. The general flow arrangement including pump locations, pump numbers and stream flow numbers is outlined in Figure 4. The average flow rate at each pumping station was determined from row five of the mass balance (Table 2). Except for biomass supply pump No. 9 which only operates 12 hours per day, the relationship between the total pounds per day values of row five and the common units of pump flow which are gallons per minute is

$$GPM = \frac{LBD}{C_3 p}$$

(9)

where: 
- GPM = flow rate (gal/min)
- LBD = mass flow rate (lb/day)
- p = weight density (lb/ft$^3$)
- $C_3$ = constant (192.5 ft$^3$-min/gal-day)

The pump flow for pump No. 9 is two times the amount predicted by Equation 9. Row one of Table 2 details the total solids in each of the twelve stream flows while row five details the total weight flow of each stream. Comparing these two rows clearly shows that at least 90 percent of all streams is simply water. Based on this fact, this analysis assumes that the density in Equation 9 is the density of water at the temperature of the liquid stream. Water temperatures are expected to range from the freezing point of 32°F to the methane phase digester operating temperature of 131°F. This temperature range provides water densities of 62.4 pounds per feet$^3$ through 61.5 pounds per feet$^3$ or approximately a 1.5 percent variation in density. The impact of this density variation is expected to be negligible and
therefore a common density of 62.0 pounds per feet³ was assumed for all flows. Equation 9 can now be simplified to

\[ GPM = \frac{LBD}{C_4} \]  

where  
\[ GPM = \text{flow rate (gal/min)} \]  
\[ LBD = \text{weight flow rate (lb/day)} \]  
\[ C_4 = \text{constant (11,935 lb-min/gal-day)} \]

Table 4 identifies the design flow rate for each pump based on Equation 10.

The other variable associated with pump sizing is total pump head. It is the effective distance in feet that the fluid mass must be raised and is equal to the difference in static head plus discharge pipe friction head plus suction pipe friction head plus velocity head. Mathematically total pump head is

\[ H = (hd - hs) + fd + fs + \frac{V^2}{2g} \]  

where  
\[ H = \text{total head (ft)} \]  
\[ hd = \text{static discharge head (ft)} \]  
\[ hs = \text{static suction head (ft)} \]  
\[ fd = \text{discharge friction head (ft)} \]  
\[ fs = \text{suction friction head (ft)} \]  
\[ \frac{V^2}{2g} = \text{velocity head (ft)} \]  
\[ V = \text{velocity of fluid in discharge pipe (ft/sec)} \]  
\[ g = \text{acceleration of gravity (32.2 ft/sec²)} \]

Equation 11 can be simplified if certain assumptions about the plant geometric configuration are made. The design criteria for piping has previously been established by limiting friction head losses to 1 psi per 100 feet of effective pipe length. Here effective
## TABLE 4

### PUMP DESIGN FLOWS

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Stream No.</th>
<th>Flow* (GPM)</th>
<th>Head (ft)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P1</td>
<td>Acid phase digester discharge - Volatile acid transfer from acid phase digester to acid surge tank and recycle spray header</td>
<td>7</td>
<td>198</td>
<td>24</td>
</tr>
<tr>
<td>P2</td>
<td>Acid storage tank discharge - Volatile acid transfer from acid surge tank to methane phase digester</td>
<td>4</td>
<td>111</td>
<td>24</td>
</tr>
<tr>
<td>P3</td>
<td>Recycle storage tank discharge - Methane phase digester effluent from recycle water tank to acid phase digester and liquid storage tank</td>
<td>5</td>
<td>110</td>
<td>24</td>
</tr>
<tr>
<td>P4</td>
<td>Acid surge tank solids - Tank bottoms to acid phase digester</td>
<td>9**</td>
<td>12**</td>
<td>24</td>
</tr>
<tr>
<td>P5</td>
<td>Liquid storage tank discharge - Plant waste effluent from liquid storage tank to the disposal site</td>
<td>10</td>
<td>91</td>
<td>24</td>
</tr>
<tr>
<td>P6</td>
<td>Supply water - Fresh water from well water supply network</td>
<td>12</td>
<td>3</td>
<td>39</td>
</tr>
<tr>
<td>P7</td>
<td>Lime slaker discharge - Lime slurry from slaker to lime slurry tank</td>
<td>3</td>
<td>3</td>
<td>24</td>
</tr>
<tr>
<td>P8</td>
<td>Lime slurry tank discharge - Lime slurry from lime slurry tank to acid surge tank and recycle water tank</td>
<td>3</td>
<td>3</td>
<td>24</td>
</tr>
<tr>
<td>P9</td>
<td>Biomass supply - Water hyacinth slurry from Lakeside harvest area to acid phase digester</td>
<td>1</td>
<td>204</td>
<td>1,880</td>
</tr>
</tbody>
</table>

* As predicted by Equation 10 from data in Table 2, row 5.
** Actual flow unknown so stream No. 9 assumed as a conservative upper limit.
pipe length is the straight run piping distance plus an allowance in equivalent straight feet for valves and fittings. Allowances for fittings are tabulated in many references like Cameron [9] and hence will not be discussed here. Near ambient temperatures, which is where this process will operate, 1 psi is equivalent to about 2.3 feet of water column. Both friction head and static head depend on the size of the plant's vessels. Since friction head is constrained to about 2.3 feet per 100 feet of effective length, it is mainly a function of the distance between tank centers. This distance depends on vessel size. Static head is the difference in elevation over which the fluid must be pumped. Hence, it also depends on vessel size. Recalling the vessel sizing criteria of 1.5 days holding time (HRT = 1.5) and inspecting Table 2 with Figure 4 shows that stream No. 4 to the methane phase digester is the largest tank influent. The methane phase digester should therefore have the largest volume (excluding the leaching-beds which have special requirements). If a cubic shape is assumed, this vessel would have a volume of about $30^3$ feet$^3$. The exact size and physical arrangement of every vessel cannot be established by a preliminary study such as this. To simplify future calculations it will be assumed that all tanks will be 30 feet tall with square cross-section. By constraining tank height to 30 feet, the active volume height of the leaching-bed will be 20 feet and Equation 7 provides a leaching-bed reactor which is about $210^2$ feet$^2$. Allowing for 50 feet between tanks, the maximum distance between tank centers is then $105 + 50 + 15$ or 170 feet. To simplify future calculations it is assumed that the maximum effective pipe length for
all pumps except No. 9 is twice the maximum tank center to center
distance or 370 feet. This value contains a generous allowance for
unknowns and provides a conservative estimate for maximum friction
head of about 8.5 feet for pumps Nos. 1 to 8. The effective length
associated with pump No. 9 is 5,300 feet. This larger value was
selected to allow the conversion facility to be located away from the
lake shore because lake frontage property is usually very expensive.
Using Equation 6 with a 5.0 feet per second velocity, a 62.0 pound per
feet$^3$ density, a 0.3 friction factor and a 0.33 feet average pipe
diameter, the friction head of pump No. 9 is about 805 psi or 1,850
feet. Inspection of Figure 4 shows that with the exception of the
biomass supply pump No. 9 and the supply water pump No. 6 all other
pumps transfer liquid from one tank to another. Since on average all
tanks will be one half full, the static pumping head for tank to tank
pumping is assumed to be 15 feet. The two unique pumps Nos. 9 and 6
are assumed to pump against the full tank height of 30 feet.

Based on the design velocities for pipings, velocity head is
expected to always be less than 1 foot. By comparison, total head is
anticipated to always exceed 20 feet. On this basis velocity head
will be neglected in this analysis because of its small contribution
to total head.

In summary, the discussion on total pump head (Equation 11) can
be condensed into the following:
1. Total friction head \((f_s + fd)\) is assumed to be 1,850 feet for pump No. 9 and for rest of the pumps total friction head is expected never to exceed 8.5 feet.

2. Total static head \((hd - hs)\) is assumed to be 30 feet for pumps No. 6 and No. 9 and 15 feet for all other pumps.

3. Velocity head \((V^2/2g)\) is assumed to be 0.

The total head predicted by these conditions is shown on Table 4. The assumptions which established these conditions are somewhat broad in scope and were selected to produce a total head estimate which is larger than what an actual plant should experience. This will produce a capital cost and pump power requirement somewhat larger than is actually required (see discussion to follow). It is assumed that this overstatement of total head will not significantly effect the results of this analysis. The validity of this assumption will be confirmed in Chapter IV's discussion of the results.

The pump driver size was selected based on the fact that power required to elevate any substance is equal to the weight flow rate times the effective lift. Mathematically this is

\[
P = w \times \text{EL}
\]

where

\[
P = \text{power (ft-lb/sec)}
\]
\[
w = \text{weight flow (lb/sec)}
\]
\[
\text{EL} = \text{effective lift (ft)}
\]

For liquids flowing in a pipe, it has been shown previously that the effective lift is equal to total head. By manipulating Equation 12 (with procedures and assumptions similar to that used in reducing
Equation 9 into Equation 10) a simplified expression for pumping power can be derived. It is

\[ WHP = \frac{GPM \times H}{C_5} \]  

(13)

where

- \( WHP \) = work horsepower (hp)
- \( GPM \) = flow rate (gal/min)
- \( H \) = total head (ft)
- \( C_5 \) = constant (3,960 gal-ft/hp-min)

Here work horsepower is the energy delivered per unit time to the fluid. The actual driver size must be somewhat larger to account for pump inefficiencies. Similarly, the actual power supplied to the pump driver must be somewhat larger to account for inefficiencies in the driver.

Inspection of Table 4 shows that all pumps, with the exception of pump No. 9, fall within the size range of 5 to 200 gallons per minute with a total head requirement of 24 to 39 feet. Review of pump manufacturing literature [12] indicates that with proper selection, pump efficiencies between 60 and 70 percent can be obtained. To simplify this analysis, a 65 percent efficiency was assumed for all pumps. This assumption implies that pump drivers should be less than 10 horsepower. Electric motors, because they provide efficiencies of 79 to 94 percent, compactness of size, and simplicity of operation are proposed for this application. For motors of less than 10 horsepower, efficiencies of 79 to 85 percent are typical [13]. To simplify this analysis, an 82 percent efficiency was assumed for all electric pump drive motors. The input power to the pump driver is then
\[ Pp = \frac{WHP}{Np \cdot Ne} \]  

where \( Pp \) = input power (hp)  
\( WHP \) = work horsepower (hp)  
\( Np \) = pump efficiency (dimensionless)  
\( Ne \) = electric motor efficiency (dimensionless)

4. **Conveyor**

Belt conveyors are used to carry water hyacinths from the lake shore to the pumping station which forwards the biomass to the conversion facility. In addition they are used to carry the residual solids out of the leaching-bed reactor. The physical size of a belt conveyor is influenced by the following factors:

1. desired material flow rate  
2. material density  
3. material shape and particle size  
4. recommended belt speed for the material  
5. developed cross sectional area of the loaded belt

The last two factors have been empirically determined by belt conveyor users and manufacturers over the years. Both are functions of the first three factors. For the application proposed, belt speeds of 400 to 600 feet per minute are recommended by Richardson [14]. The developed cross sectional area describes the shape of load as it sits on the moving belt. Multiplying belt speed times the load's cross sectional area establishes the conveyor's transfer rate in feet\(^3\) per minute. As stated previously, the cross sectional area is an empirical factor and it is different for each belt width. These factors are
<table>
<thead>
<tr>
<th>Belt Width (in)</th>
<th>Load's Cross Sectional Area (ft) [14]</th>
</tr>
</thead>
<tbody>
<tr>
<td>18</td>
<td>.18</td>
</tr>
<tr>
<td>24</td>
<td>.333</td>
</tr>
<tr>
<td>30</td>
<td>.533</td>
</tr>
<tr>
<td>36</td>
<td>.780</td>
</tr>
<tr>
<td>42</td>
<td>1.100</td>
</tr>
</tbody>
</table>

The weight transfer rate of a belt conveyor is the product of the load's cross sectional area, belt speed and weight density. Mathematically weight transfer rate is

\[
w = XS \times V \times p
\]  

where \( w \) = weight transfer rate (lb/sec)  
\( XS \) = load’s cross sectional area (ft\(^2\))  
\( V \) = belt speed (ft/sec)  
\( p \) = weight density (lb/ft\(^3\))

The common units of belt conveyor capacity are ton per hour. Equation 15 can be manipulated into this form by the addition of constants and this more useful form is

\[
TH = XS \times V \times p \times C_6
\]  

where \( TH \) = belt conveyor capacity (ton/hr)  
\( C_6 \) = constant (1.8 ton-sec/hr-lb)

The power to drive a belt conveyor is composed of three components. They are as follows:

1. Power to keep an empty belt moving. This results from the work of bending the belt around the end pulleys and from frictional work associated with belt guides.

2. Power to move the load horizontally. This results from frictional work associated with the load's weight acting on
the belt guides and from accelerating the material up to the belt speed.

3. Power to move the load vertically. This results from raising the load against the force of gravity.

Items (1) and (2) above have been empirically determined by users and manufacturers over the years. Item (3) is simply an application of the definition of power (see Equation 12). The following relations express the power requirements of these components. Specifically the power to move an empty belt is the product of an empirical belt flexing constraint, belt speed and conveyor length or

$$Pe = CB(wd) \times S \times l$$

(17)

where

- $Pe =$ shaft power to move empty belt (hp)
- $CB(wd) =$ empirical belt flexing power constraint [13] (hp-min/ft²)
- $wd =$ belt width in inches
- $CB(18) = .000020$
- $CB(24) = .000026$
- $CB(30) = .000036$
- $CB(36) = .000047$
- $CB(42) = .000053$
- $S =$ belt speed (ft/min)
- $l =$ conveyor length (ft)

Here conveyor length is the same as the collection boom diameter (to be discussed shortly) or the leaching-bed width.

The shaft power to move the load horizontally is the product of an empirical function of length and weight flow in tons per hour. Mathematically this relation is

$$Ph = (.004 + .0000325 \times l) \times TH$$

(18)

where

- $Ph =$ shaft power to move load horizontally (hp) [13]
- $l =$ conveyor length (ft)
- $TH =$ conveyor capacity (ton/hr)
The shaft power to move the load vertically is the product of weight flow and effective lift or

\[ P_v = TH \times EL \times C_7 \]  

(19)

where  
- \( P_v \) = shaft power to lift load vertically (hp)  
- \( TH \) = conveyor capacity (ton/hr)  
- \( EL \) = effective lift (ft)  
- \( C_7 \) = constant (.001 hp-hr/ton-ft)

The belt conveyer driver size is the sum of the three components or

\[ P_t = P_e + P_h + P_v \]  

(20)

where  
- \( P_t \) = total shaft power (hp)

For the same reasons outlined for pumps, an electric motor drive with efficiency \( Ne = .82 \) is assumed for the belt conveyor. The input power is then

\[ P_c = \frac{P_t}{Ne} \]  

(21)

where  
- \( P_c \) = input power (hp)  
- \( Ne \) = electric motor efficiency (dimensionless)

5. **Winches**

Winches are motor driven machines which provide motive force to loads by winding a flexible cable around a drum. They are required in this system to drag water hyacinths from their growing location to the lake shore (see Figure 6). Once a day, the winches will be used to drag a collection boom to the water hyacinth growing area. A day’s feed supply will be encircled and then the winches will be used to drag the plants to the shore. At shore, the winches will hold the
plants in position until they can be picked up and conveyed to the plant. This cycle must be repeated each day as the conversion facility requires continuous feed. To establish winch sizes which would provide the daily feedstock requirements it was necessary to assume the following operating schedule:

1. 3 hours to position boom
2. 9 hours to bring plants to shore
3. 12 hours to hold plants at shore while they are removed

The speed at which the winch cable is withdrawn and the force which it must pull against, establish the size and power requirements of the winches. The speed of travel is determined by the longest dimension of the farm and the travel time assumed by the operating schedule.

Farm size is established by the growth productivity of the water hyacinth plants and the feed requirements of the conversion facility. The productivity of water hyacinths ranges from 8 to 52 dry tons per acre-year [15]. This study assumes an average value of 33.3 dry tons per acre-year as suggested by Reddy [16] because this value can easily be obtained in Florida with moderate applications of fertilizer. The feed requirements of the conversion facility are obtained by applying the unit process conversion factors of

1. fraction volatile solids of dry feedstock matter
2. conversion rate of volatile solids to gas
3. BTU content of gas

to the design plant output. The unit conversion factors for water hyacinths are shown in Table 3. In summary, the yearly feedstock requirement is
\[
DTY = \frac{X}{hc \times CR \times FVS \times C_8}
\]

where

\( DTY \) = yearly feedstock requirement (dry ton/yr)
\( X \) = yearly energy production of facility (BTU/yr)
\( hc \) = BTU content of gas (1,000 BTU/ft\(^3\))
\( CR \) = conversion rate (SCF/lb volatile solids)
\( FVS \) = fraction volatile solids (dimensionless)
\( C_8 \) = constant (2,000 lb/ton)

For a production level of \(0.1 \times 10^{12}\) BTU per year, Equation 22 implies that about 9,690 dry tons per year will be required by the plant. With a yield of 33.3 dry tons per acre-year the farm would require about 291 acres. The conceptual design of the farm in Figure 6 shows a triangular shape. This was assumed to minimize the number of winches. With a configuration of base equal to height for simplicity, the 291 acre farm will have a 5,035 feet base and height. To traverse this distance in 3 hours requires a 0.47 feet per second velocity while 9 hours implies a 0.16 feet per second velocity.

The force on the winch cables result from four forces associated with moving the water hyacinth mats through the environment. To simplify the analysis it is assumed in the discussion which follows that all pulling forces act in parallel with the load forces. This is a reasonable assumption if the winches are located a distance apart which is equal to the diameter of the collection boom and if the winch elevation is the same as the water level of the lake. The four forces acting on the towed water hyacinths are as follows:

1. \textit{Form drag due to air} - This is the force which results from the normal impact of a body’s frontal area with the atmosphere.
2. Friction drag due to air - This is the force which results from tangential friction between a body's top surface and the atmosphere.

3. Form drag due to water - This is the force which results from the normal impact of a floating body's frontal area with the water.

4. Friction drag due to water - This is the force which results from tangential friction between a floating body's bottom surface and the water.

All four of these forces can be predicted by the general drag equation [13]. This equation (23) relates drag force to the relative fluid velocity and to the characteristic area through an empirical constant which must be determined experimentally and depends on Reynolds number. The general drag equation is

\[ DF = \frac{Cd \cdot p \cdot V^2 \cdot A}{2g} \]  

(23)

where
- \( DF \) = drag force (lb)
- \( Cd \) = empirical drag coefficient (dimensionless)
- \( p \) = weight density (lb/ft\(^2\))
- \( V \) = relative velocity (ft/sec)
- \( A \) = characteristic area (ft\(^2\))
- \( g \) = acceleration of gravity (32.2 ft/sec\(^2\))

Characteristic area is the body's normal frontal area for form drag and the plane surface area for friction drag. Both of these areas result from shape of the water hyacinth mass as it is towed in the water. The area of water hyacinths which must be harvested each
day is the daily feedstock requirement (Table 2, column 1, row 1) divided by standing plant density, fraction dry matter and the number of pounds per ton (Table 3). Mathematically this is

\[ HA = \frac{DTD}{SD \cdot FDM} \] (24)

where
- \( HA \) = harvest area (acre/day)
- \( DTD \) = daily feedstock requirement (dry ton/day)
- \( SD \) = standing plant density (wet ton/acre) Table 3
- \( FDM \) = fraction dry matter (dimensionless)

For the 0.1(10E12) BTU per year production facility Equation 24 predicts that 3.66 acres must be harvested each day. Since the towing boom is flexible, it is expected that the plant mass will take a circular shape as it is towed. To simplify the analysis, a half-circle shape is assumed. With this simplification, the normal area of the body is the length above water or root length times the half-circle diameter. Normal area is

\[ An = D \cdot 1 \] (25)

where
- \( An \) = normal area (ft\(^2\))
- \( D \) = diameter of half-circle (ft) \(= (HA \cdot C_9 \cdot 8/\pi)^{0.5} \)
- \( C_9 \) = constant \((43,560 \text{ ft}^2\text{-day/acre})\)
- \( l \) = length of root or top (ft)

The tangential friction area is simply the half-circle area or

\[ At = HA \cdot C_9 \] (26)

where
- \( At \) = tangential friction area (ft\(^2\))
- \( HA \) = harvest area (acre/day)
- \( C_9 \) = constant \((43,560 \text{ ft}^2\text{-day/acre})\)
For a predicted harvest area of 3.66 acres, Equations 25 and 26 estimate a normal root area of 960 feet\(^2\), a normal top area of 1,590 feet\(^2\) and a tangential area of 159,600 feet\(^2\) respectively.

The drag coefficients (Cd) are empirical and must be determined by experiment. Values for flat plates and blunt objects can be found in standard references [13]. However, the irregular shape of the water hyacinth plants makes their direct application unsound. The U.S. Army Corps of Engineers performed tests on pushing water hyacinth masses with motor boats [17]. Tests were conducted in open river water which had a current of .25 feet per second and a wind speed of 0.0 feet per second. A 314 feet\(^2\) area of water hyacinth mat was surrounded with a flexible boom and pushed up and back a 600 foot test course. Force readings between the pushing rake and the boat were taken every 15 seconds and averaged. The test course was traversed four times at constant ground speeds between 0.75 and 2.50 feet per second. The results of this test along with the physical test conditions are detailed in Table 5. To correlate this data with the standard drag coefficients, Table 6 was prepared using Equation 23 with the assumption that the mat of water hyacinths would act like a blunt flat plate. Each of the four force element's contribution to the total is shown. Inspection of predicted forces indicates that form drag always dominates friction drag and that the predictions are about four times larger than the actual test results. This is reasonable because the investigators observed that the plant bodies bend to form a smooth surface as they are moved through the water [17].
TABLE 5
TEST DATA - WATER HYACINTH MAT PUSHING TEST NO. 3

<table>
<thead>
<tr>
<th>Against Flow</th>
<th>With Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ground Speed (ft/sec)</td>
<td>Force (lb)</td>
</tr>
<tr>
<td>0.76</td>
<td>21</td>
</tr>
<tr>
<td>1.28</td>
<td>40</td>
</tr>
<tr>
<td>1.85</td>
<td>72</td>
</tr>
<tr>
<td>2.08</td>
<td>100</td>
</tr>
</tbody>
</table>

SOURCE: P. A. Smith, Mechanical Harvesting of Aquatic Plants, Report 2, Environmental Laboratory, U. S. Army Engineer Waterways Experiment Station, Vicksburg, Mississippi, 1980.

NOTES: Test conditions
- Surface area: 314 ft²
- Shape of mass: round
- Diameter: 20 ft
- Top length: 2.67 ft
- Root length: 1.85 ft
- Wind speed: 0 ft/sec
- Current speed: 0.25 ft/sec
TABLE 6
PREDICTED FORCES FOR PUSHING A CIRCULAR 314 SQUARE FOOT WATER HYACINTH MAT

<table>
<thead>
<tr>
<th>Ground Speed (ft/sec)</th>
<th>Relative</th>
<th>Air Velocity (ft/sec)</th>
<th>Water Velocity (ft/sec)</th>
<th>Force Elements* DF₁</th>
<th>DF₂</th>
<th>DF₃</th>
<th>DF₄</th>
<th>Total Force (lb)</th>
<th>Ratio Act./Est. (lb)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
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<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Against Flow</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>.76</td>
<td>0.76</td>
<td>1.01</td>
<td>.07</td>
<td>.0015</td>
<td>73</td>
<td>1.24</td>
<td>74</td>
<td>.28</td>
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<tr>
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<td>.20</td>
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<tr>
<td>1.85</td>
<td>1.85</td>
<td>2.10</td>
<td>.43</td>
<td>.0089</td>
<td>316</td>
<td>5.36</td>
<td>322</td>
<td>.22</td>
<td></td>
</tr>
<tr>
<td>2.08</td>
<td>2.08</td>
<td>2.33</td>
<td>.55</td>
<td>.0112</td>
<td>389</td>
<td>6.60</td>
<td>396</td>
<td>.25</td>
<td></td>
</tr>
<tr>
<td><strong>With Flow</strong></td>
<td></td>
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<td></td>
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<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.25</td>
<td>1.25</td>
<td>1.00</td>
<td>.20</td>
<td>.0041</td>
<td>72</td>
<td>1.22</td>
<td>72</td>
<td>.28</td>
<td></td>
</tr>
<tr>
<td>1.72</td>
<td>1.72</td>
<td>1.47</td>
<td>.37</td>
<td>.0077</td>
<td>155</td>
<td>2.63</td>
<td>158</td>
<td>.26</td>
<td></td>
</tr>
<tr>
<td>2.27</td>
<td>2.27</td>
<td>2.02</td>
<td>.65</td>
<td>.0134</td>
<td>292</td>
<td>4.96</td>
<td>298</td>
<td>.20</td>
<td></td>
</tr>
<tr>
<td>2.42</td>
<td>2.42</td>
<td>2.17</td>
<td>.74</td>
<td>.0152</td>
<td>337</td>
<td>5.72</td>
<td>344</td>
<td>.26</td>
<td></td>
</tr>
<tr>
<td><strong>Average of all values</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>.245</td>
</tr>
</tbody>
</table>

NOTES:
- Mat diameter ............... 20 ft
- Top length .................. 2.67 ft
- Root length ................. 1.85 ft
- Wind speed .................. 0.0 ft/sec
- Current speed .............. 25 ft/sec
- Water weight density ...... 62.4 lb/ft³
- Air weight density ....... 0.076 lb/ft³
- $C_d_1$ ..................... 2.0
- $C_d_2$ ..................... 0.007
- $C_d_3$ ..................... 2.0
- $C_d_4$ ..................... 0.004

* Per Equation 23.
the plant bodies are a collection of appendages the formation of the smooth surface would have the effect of reducing the mat's normal area. The average ratio of actual measured force to estimated force is 0.245. Based on this fact future calculations will assume that form drag of a water hyacinth mat is equal to 25 percent of the form drag of a blunt-flat plate. In addition, since friction drag is less than 2 percent of the form drag, it will be neglected in this analysis.

The relative velocity of a body moving in a fluid is the vector sum of the body velocity and the fluids velocity both with respect to the same coordinate system. For air, the relative velocity is the sum of the motion toward shore plus wind velocity. There are two conditions of wind velocity of interest. The first is expected maximum velocity while the second is average resultant velocity and its direction. Maximum velocity establishes the largest force to be expected and hence, is used to determine winch size. Average resultant velocity is the vector sum of all wind velocity measurements divided by the number of observations. It establishes the average expected force vector and is used to calculate the winches' yearly energy consumption from the definition of work. In the discussion that follows, it is assumed that both the wind and current vectors act in a plane parallel to the surface of the water. This is necessary to simplify the analysis. The maximum wind velocity is assumed to be 94 feet per second while the average resultant velocity is assumed to be 1.6 feet per second at 90°. These values have been recorded for the Central Florida city of Orlando [18]. Towing speed toward shore was previously
shown to be considerably less than 1 foot per second. Based on these two facts, the design analysis will use a value of 94 feet per second for the maximum relative velocity of air. Since the exact orientation of the hyacinth farm is unknown, this analysis will assume the resultant velocity vector opposes the mat's movement toward shore. To simplify future calculations, a 2.0 feet per second average relative velocity for air will be used in the energy consumption calculations. Determining the magnitudes of water currents in Florida's lakes goes beyond this study's level of detail. It is expected that the average resultant water current will be zero in a confined body of water like a lake. Since better data is not readily available, twice the observed river current [17] or 0.5 feet per second will be assumed for the maximum water current velocity. Based on the above the design analysis will use 1.0 feet per second as the relative water velocity and the energy analysis will use 0.5 feet per second as the relative water velocity. The total force exerted by the plant mass on the winches is

\[
DFT = DF_1 + DF_2 + DF_3 + DF_4
\]  

(27)

where

- \( DFT \) = total drag force on winches (lb)
- \( DF_1 \) = form drag force of air (lb) per Equation 23
- \( DF_2 \) = friction drag force of air (lb) per Equation 23
- \( DF_3 \) = form drag force of water (lb) per Equation 23
- \( DF_4 \) = friction drag force of water (lb) per Equation 23

The driver size for the winches is established from the definition of power. Power is the dot product of force and velocity. Cable force is given by Equation 27. Since cable velocity is coincident with
force, power reduces to the simple product of cable velocity and cable force. The power to the cables is

\[ P = DFT \times V \]  

(28)

where \( P \) = power (ft-lb/sec)  
\( DFT \) = total drag force (lb)  
\( V \) = cable velocity (ft/sec)

Mechanical efficiencies for winches are about 80 percent [13]. To maintain consistency an electric motor efficiency of 82 percent is assumed. The input power to the winches is then

\[ P_w = \frac{DFT \times V}{Nw \times Ne \times C_{10}} \]  

(29)

where \( P_w \) = input power (hp)  
\( DFT \) = total drag force (lb)  
\( V \) = cable velocity (ft/sec)  
\( Nw \) = winch efficiency (dimensionless)  
\( Ne \) = electric motor efficiency (dimensionless)  
\( C_{10} \) = constant (550 ft-lb/sec-hp)

6. Gas Cleaning and Compression

The gas produced by the methane phase digester is expected to be 85 percent methane and 15 percent carbon dioxide by volume and at atmospheric pressure [4]. Before this gas can be introduced into the existing natural gas distribution system its carbon dioxide component must be lowered to 3 percent and its pressure must be raised to the nominal distribution pressure of 225 psig [2]. The gas clean-up process utilized in this analysis is the 'Kryosol Process' by Kryos Energy, Inc. This patented (No. 4,252,548) process uses a chilled methane absorbant. Since this is a proprietary process, few details of the process are available to the public. The following description represents the author's understanding of the process. The methane and
carbon dioxide mixture is first compressed to 10 psi above the desired outlet pressure. The compressed gas mixture is then forced through chilled methanol in an absorption tower where the carbon dioxide is absorbed. The absorbant is then moved to a low pressure regenerator where the carbon dioxide is allowed to boil off into the atmosphere. This boiling action provides all the refrigeration necessary for the 'Kryosol Process'. Finally, the regenerated and chilled absorbant is pumped back to the absorption tower so the cleaning cycle can be repeated. Pumping power can be supplied by expanding the waste carbon dioxide through a back pressure turbine before discharge to the atmosphere. The 'Kryosol Process' was selected for this study because no energy inputs (other than the 10 psi pressure drop) are required, it is commercially available in the required size range and it has been successfully applied in a similar application like land fill gas containing methane. Individual components of this system are sized by the supplier. The only sizing information required is the inlet gas flow rate provided by column 6, rows 7 and 8 of Table 2 and the inlet pressure of 235 psig.

The power requirements of the gas compressor can be determined from the general energy equation for fluids or

\[
\frac{V_1^2}{2g} + ps_1v_1 + u_1 + Z_1 + W + Q = \frac{V_2^2}{2g} + ps_2v_2 + u_2 + Z_2
\]  

(30)

where

- \( V \) = fluid velocity (ft/sec)
- \( g \) = gravitation constant (32.2 ft/sec²)
- \( ps \) = static pressure, absolute (lb/ft²)
- \( v \) = specific volume (ft³/lb)
- \( u \) = internal energy (ft-lb/lb)
- \( Z \) = elevation above datum (ft)
\[ W = \text{work done on fluid (ft-lb/lb)} \]
\[ Q = \text{heat added to fluid (ft-lb/lb)} \]

For the application proposed, the specific volume/static pressure product is 41,000 feet for the inlet and 76,000 feet for the outlet condition. Note that the design piping velocity is less than 215 feet per second and that most compressors are less than 10 feet in height. Based on these two facts, both elevation head and velocity head are insignificant when compared to static head. Since continuous compression is required by the process, it is expected that compression will be adiabatic like most commercial compressors [19]. This analysis neglects velocity head, elevation head and heat exchange. With these simplifications Equation 30 reduces to

\[ W = (p_s_2 v_2 + u_2) - (p_s_1 v_1 + u_1) = h_2 - h_1 \]  \hspace{1cm} (31)

By definition, \( pv + u \) is enthalpy and hence, the right side of Equation 31 is an enthalpy difference. The specific heat for most gases near ambient temperature is nearly constant. With this assumption the difference in enthalpy is approximately equal to the product of specific heat at constant pressure times the change in temperature. Assuming ideal gas behavior for this isentropic process, it can be shown that temperature and pressure are related by

\[ \frac{(p_s_2)}{(p_s_1)} \left( \frac{(K-1)}{K} \right) = \frac{T_2}{T_1} \]  \hspace{1cm} (32)

where \( p_s = \text{static pressure, absolute (lb/ft}^2) \)
\( T = \text{temperature (R)} \)
\( K = \text{ratio of specific heats (dimensionless)} \)

Substituting Equation 32 into Equation 31 along with the approximation for enthalpy difference and noting that specific heat at constant
pressure minus the specific heat at constant volume equals the ideal gas constant provides a simplified expression for the unit adiabatic work to compress one pound mole of an ideal gas from \( P_1 \) to \( P_2 \). Mathematically this relationship is

\[
W' = \frac{K R T_1 \times ((ps_2 / ps_1)^{((K-1)/K)} - 1)}{(K-1)}
\]  

(33)

where \( W' = \) work done on fluid (ft-lb/lb mole)
\( K = \) ratio of specific heats (dimensionless)
\( R = \) ideal gas constant (1,544 ft-lb/lb mole-R)
\( T_1 = \) initial fluid temperature (R)
\( ps_1 = \) initial fluid static pressure, absolute (lb/ft\(^2\))
\( ps_2 = \) final fluid static pressure, absolute (lb/ft\(^2\))

Equation 33 predicts the energy required for compression. The actual drive must be somewhat larger to account for pressure drop across valves and for mechanical friction within the compressor. Typical overall compressor efficiencies range from 65 to 82 percent [13]. This analysis assumes an overall efficiency of 73.5 percent. This average value was selected because detailed study of this one area was not practical. A rough approximation for driver size indicates that in excess of 100 horsepower will be required. Typical electric motor efficiency for this size motor is 91 percent. Previously an electric motor efficiency of 82 percent was utilized. To simplify the analysis and to maintain consistency, 82 percent will be used in the compressor analysis. This assumption is reasonable because it partially offsets the unknowns associated with the 73.5 percent compressor efficiency. With these assumptions the input power to the compressor is
\[
P_x = \frac{K R T_1 m}{(K - 1) \text{Ne} \text{Nx} C_{10}} \left(\frac{p_{s2}}{p_{s1}}\right)^{(K-1)/K} - 1 \tag{34}
\]

where
- \(P_x\) = input power to compressors (hp)
- \(K\) = ratio of specific heats (dimensionless)
- \(R\) = ideal gas constant (1,544 ft-lb/lb mole-R)
- \(T_1\) = inlet temperature (R)
- \(m\) = mass flow rate (lb mole/sec)
- \(p_{s2}\) = final absolute static pressure (lbs/ft\(^2\))
- \(p_{s1}\) = initial absolute static pressure (lbs/ft\(^2\))
- \(\text{Ne}\) = electric motor efficiency (dimensionless)
- \(\text{Nx}\) = compressor efficiency (dimensionless)
- \(C_{10}\) = constant (550 ft-lb/sec-hp)

**Energy Balance Considerations**

An economically viable and technically feasible process must produce more energy than it consumes. This analysis investigates the energy effectiveness of the proposed process by examining an overall energy balance. The general equation for energy balance is

\[
\text{Accumulation of energy within the system} = \text{Transfer of energy into system through system boundary} - \text{Transfer of energy out of system through system boundary} + \text{Energy generation within system} - \text{Energy consumption within system} \tag{35}
\]

Here the system is composed of the farm harvesting system and the conversion plant as shown in Figure 7. The system boundary is the imaginary edge formed by the union of the system's two components. The purpose of the discussion which follows is to develop a methodology to determine if energy production exceeds energy consumption.
Figure 7. System Definition For Energy Balance
Each of the terms in Equation 35 has a specific meaning when analyzing the proposed system. Accumulation of energy refers to the storage of energy within the system. Since the proposed process is continuous, this term is equal to zero. Transfer of energy into the system involves three components. They are electrical energy which does work in the process, thermal energy to maintain the process at operating temperature and the chemical potential energy contained in the biomass feedstock. Transfer of energy out of the system also involves two components. They are the BTU content of the gas produced and thermal heat losses to the environment. Energy generation results from chemical reactions in the process. The overall effect of all chemical reactions is exothermic. For the proposed system, it is assumed that heat generated by the exothermic chemical reactions is sufficient to offset heat losses to the environment. This assumption is based on the observations of Colleran [4]. Energy consumption within the system results from mechanical work in the process. Since electricity is only used to power electric motors, it is assumed that mechanical work plus some small thermal losses due to inefficiencies are equal to the total transfer of electrical energy into the system.

The purpose of this investigation is to establish whether net energy production occurs. To answer this question it is not necessary to perform a detailed energy balance. The preceding discussion has limited the question of net energy production to whether the energy content of methane produced exceeds the electric energy used for work in the process. This results from the fact that heat losses are
supplied by the overall exothermic chemical reactions, there is no energy accumulation within the process, and except for conversion inefficiencies, all electrical energy inputs are utilized to do work in the process. Mathematically the net energy question can be simplified into

\[
\begin{bmatrix}
\text{Energy content of gas produced} \\
\end{bmatrix} \quad ? \quad \begin{bmatrix}
\text{Electrical energy required to produce gas} \\
\end{bmatrix}
\] (36)

Mechanical work of the proposed process can be classified into several categories. They are as follows:

1. pump work to move fluids
2. conveyor work to move solid material
3. towing work to harvest feedstock material
4. grinding work to densify material prior to slurry preparation
5. compression work to raise gas to pipeline pressure

The energy consumption of pumps and conveyors is the product of their power requirement times operating time. For pumps Equation 14 predicts power requirements. Since the proposed process is assumed to be continuous, operating time for pumps will be 24 hours per day, 365 days per year, except for the biomass supply pump which operates simultaneously with the conveyors. For conveyors Equation 21 predicts power requirements. Based on the assumed operating cycle of the harvesting equipment, operating time of the shore conveyers will be 12 hours per day, 365 days per year. Since the exact operating cycle of the residual solids' conveyor is undefined, estimating operating hours is impractical. Energy consumption for these conveyors was estimated
as a fraction of the shore conveyor's consumption. The fraction used for this analysis is the ratio of weights handled as specified by row No. 4, columns Nos. 1 and 9 of Table 2. In other words, 11 percent of shore conveyor consumption was allowed for residual solid conveyor consumption. Energy consumption which results from towing will be calculated from the definition of work. Work is the dot product of force and distance. Since cable force is coincident with cable travel, work reduces to the simple force-distance product. Here force results from moving the water hyacinths on the surface of the water and is predicted by Equation 27. Distance in terms of energy consumption, is twice the average distance between the shore conveyors at one corner and the center of the plot. For the proposed farm arrangement where base equals height, average distance (which must be traversed once each day) is equal to the base length times the square root of 2. This results from the geometry of right triangles. The efficiencies associated with the towing system are 80 percent for the winches and 82 percent for the electric motor drive [13]. In summary, the yearly energy consumption for pumps, conveyors and towing are, respectively

\[ E_p = P_p \times t \]  
\[ E_c = P_c \times t \]  
\[ E_w = \frac{DFT \times TC \times l \times C_{11}}{N_w \times N_e \times C_{10} \times C_{12}} \]  

where

- \( E_p \) = yearly energy consumption of pumps (hp-hr/yr)
- \( E_c \) = yearly energy consumption of conveyors (hp-hr/yr)
- \( E_w \) = yearly energy consumption of winches (hp-hr/yr)
- \( P_p \) = input power to pumps (hp) Equation 14
- \( P_c \) = input power to conveyors (hp) Equation 21
- \( t \) = operating time per year (hr/yr)
DFT = total drag force on winches (lb)
TC = towing constant for triangle farm (1.41 ft/day-ft)
1 = length of triangle base (ft)
C_{11} = constant (365 days/yr)
Nw = winch efficiency (dimensionless)
Ne = electric motor efficiency (dimensionless)
C_{10} = constant (550 ft-lb/sec-hp)
C_{12} = constant (3,600 sec/hr)

Grinding energy is utilized to reduce whole water hyacinth plants to nominal 0.5 inch pieces. This is necessary to form the slurry which will be pumped to the conversion facility in a pipeline as well as for rapid processing once the material reaches the conversion facility. The energy to grind any material depends on initial particle size and internal structure. Grinding energy is an empirical constant with units of horsepower-hour per ton and must be determined by experiment. For water hyacinths, little experimental work has been attempted on grinding but Bagnall obtained a grinding energy of about 120 horsepower-hour per wet ton using a shear-bar type chopper [20]. For lack of better data, this study will utilize this value and hence, the yearly energy consumption for grinding is

\[ E_g = \frac{WG \times DTY}{FDM} \]  \hspace{1cm} (40)

where \( E_g \) = yearly energy consumption of grinder (hp-hr/yr)  
\( WG \) = grinding work constant (hp-hr/green ton)  
\( DTY \) = yearly feedstock requirement (dry ton/yr)  
\( FDM \) = fraction dry matter (dimensionless)

Compression work results from raising gas pressure from 14.7 psia as it leaves the methane phase digester to 250 psia which is the
required inlet pressure of the gas cleaning equipment. The energy consumption of the compressor is the product of their power requirement times operating time. Power requirements for compressors are given by Equation 34. Since the proposed process is assumed to be continuous, operating time will be 24 hours per day, 365 days per year. The yearly energy consumption by compressors is then

\[ Ex = Px \times t \]  

(41)

where \( Ex \) = yearly energy consumption of compressors (hp-hr/yr)  
\( Px \) = input power to compressor (hp)  
\( t \) = operating time per year (hr/yr)

As stated previously, the objective of this energy analysis is to determine if the BTU content of the produced gas exceeds the electrical energy consumed to produce the gas (see Equation 36). The electric energy to produce the yearly gas output of the conversion facility is the sum of the yearly energy consumption of pumps, conveyors, winches, grinders and compressors. With this in mind Equation 36 can be recast into

\[ X > (Ep + Ec + Ew + Eg + Ex) \times C_{13} \]  

or

\[ X > E \]

(42)

where \( X \) = plant BTU output per year (BTU/yr)  
\( Ep \) = yearly energy consumption of pumps (hp-hr/yr)  
\( Ec \) = yearly energy consumption of conveyors (hp-hr/yr)  
\( Ew \) = yearly energy consumption of winches (hp-hr/yr)  
\( Eg \) = yearly energy consumption of grinders (hp-hr/yr)  
\( Ex \) = yearly energy consumption of compressors (hp-hr/yr)  
\( C_{13} \) = constant (2,545 BTU/hp-hr)  
\( E \) = yearly energy consumption of electricity (BTU/yr)
**Economic Assumptions**

Economic analysis of any proposed process is necessary to establish whether the concept will be implemented through natural economic forces, to determine which components are the significant cost contributors and to determine if any design assumptions significantly effect the economic viability of the process. A life cycle cost methodology will be used for this analysis. This technique compares the initial capital costs to the present value of the lifetime cash flow for various assumed economic lives. Cash flow includes operating costs and gas sale revenues. The economic life which produces a cash flow present worth equal to the initial capital is the payback period. If the payback period is less than the proposed equipment life, then the process is economically viable. The specifics of this methodology will be discussed shortly. In the discussions on capital costs and operating costs which follow, all values are in 1981 dollars. This assumption was necessary because complete cost data for more recent years is not yet available to the author.

Capital costs were established by first determining the size of major components by the methods previously outlined in this chapter. Next the direct cost of labor and materials for these major components were estimated by either consulting a standard cost estimating reference [14] or by obtaining vendor quotations. The majority of cost data in this analysis was obtained from references because most of the equipment involved is conventional process equipment and has
well documented costs. Only manufacturers of specialty items like gas
clean-up equipment which could not be readily estimated through
standard references were contacted for quotations. Since this is a
limited scope study, instrumental and electrical costs were not
considered in detail. Instead a flat 10 percent of non-site related
direct costs was allowed for each. This is a typical value used for
preliminary studies of this type [21]. Direct costs cover the price
of material and labor to install equipment. Contingency, contractor
overhead/profit, engineering design and spare parts are indirect costs
which must be added to direct costs to obtain total construction
costs. Indirect costs are generally estimated as a percentage of
direct costs and these percentages can also be found in standard
references [22]. This analysis utilizes the following values:

1. contingency 20%
2. contractor overhead/profit 30%
3. engineering design 9%
4. spare parts 1.6%

Operating and maintenance costs occur from the four generic
sources. They are as follows:

1. personnel
2. maintenance materials
3. chemicals
4. electricity

Personnel costs were established by identifying the labor functions in
the proposed facility, estimating the number of people required to
accomplish these functions and finally applying an hourly wage rate to
the total man-hours associated with the number of required people. Table 7 outlines the tasks and the number of people required for the proposed process. A total of 20 people paid for 40 hours per week, 52 weeks per year will be required. To simplify the analysis a single average loaded hourly rate was used for all personnel. A loaded rate includes the dollar amount of the wages which will appear in a worker's paycheck as well as an allowance for the cost of benefits like social security, workmen's compensation and medical insurance. This analysis assumes the allowance for benefits is 50 percent of direct wages for simplicity [23]. The average direct wage was assumed to be $9.33 per hour because this value is typical of the Central Florida citrus industry and because it provides and average loaded hourly rate of $14.00 per hour [24]. Maintenance material costs were estimated as a percentage of construction costs. Although no references could be found on this subject, it is reasonable to expect that maintenance repairs over the life of the plant will be about equal to the original construction costs. In other words, on average everything in the plant will require replacement during the plant's life. The expected life of this plant is 20 to 40 years. This expected life range suggests an annual maintenance repair parts expense of 2.5 to 5 percent of construction dollars. Since a large portion of the plant's equipment is non-moving part devices like tanks, pipes, etc., a 2.7 percentage (one near the low end of the range) was assumed reasonable for this application. Chemical expenses
<table>
<thead>
<tr>
<th>Function</th>
<th>Number of People</th>
</tr>
</thead>
<tbody>
<tr>
<td>Plant manager</td>
<td>1</td>
</tr>
<tr>
<td>Plant engineer</td>
<td>1</td>
</tr>
<tr>
<td>Quality control chemist</td>
<td>1</td>
</tr>
<tr>
<td>Secretary/bookkeeper</td>
<td>1</td>
</tr>
<tr>
<td>Operating technician</td>
<td>4</td>
</tr>
<tr>
<td>(1 per shift plus relief)</td>
<td></td>
</tr>
<tr>
<td>Maintenance/utility worker</td>
<td>10</td>
</tr>
<tr>
<td>Farm worker</td>
<td>2</td>
</tr>
<tr>
<td>(1 for 2 shifts per day, relief from utility)</td>
<td></td>
</tr>
</tbody>
</table>

**Total** 20
include fertilizer for the farm and lime for the conversion facility. Lime requirements are detailed in column 5, row 1 of the mass balance Table 2. The cost of delivered lime is assumed to be $71 per ton [25]. Fertilization expenses were not investigated in detail. Instead a value of $0.50 per dry ton water hyacinth was assumed as suggested by Reddy [16]. Electricity costs were estimated by applying electricity's unit cost to the total yearly consumption estimated by the energy analysis (Equation 42). The unit cost of electricity is assumed to be $13.50 per million BTU. This value is typical of rates charged to industrial customers in Central Florida [26].

The revenues from gas sales are assigned a dollar value of $6.00 per million BTU [27]. This amount is the current value used by the Gas Research Institute for Evaluating Emerging Technologies. It represents the gas industry's expectations of the wholesale price in 1981 dollars of natural gas in the year 2000.

This study's life cycle cost analysis utilizes a present worth technique. Present worth is the lump sum value of a series of payments. In a capitalistic society like the United States, there is a cost associated with the use of money and hence, capital. This cost is commonly referred to as interest and is the money paid for the use of borrowed money. The ratio of the money paid to the money borrowed per unit time is called interest rate. For capital projects like the proposed project, it is customary for interest to be paid yearly. The nature of interest and borrowed money which is important to any
economic analysis is that up front expenditures are always less than the repayment amount. Hence, the present worth of future payments is always less than the sum of those payments and a dollar in today's economy has a larger value than a dollar in tomorrow's economy. This time value of money requires that economic alternatives be compared in same year dollars. For the proposed process the economic alternatives are to purchase gas from another supplier at $6.00 per million BTU or construct and operate a biomass to methane facility of finite life. If the present value of purchasing gas over a period of time exceeds the present value of construction and operation of the facility over the same time period, then methane from biomass is an economically sound idea. To judge how attractive an economic alternative is, it is common to determine how many years of operation are required to pay off an investment. This is accomplished by comparing the present value of the two alternatives for different repayment periods. The time to pay off the investment results when the two present worths are equal in magnitude. If the payoff time is approximately the same or exceeds the equipment life, then the alternative is not a sound investment. If the payoff time is considerably less than the equipment life, then the alternative is an attractive investment. The alternative becomes more attractive as the difference between payoff time and equipment life increases.

Assuming annual interest, the value of one dollar in one year is $(1 + i)$ where $i$ is the interest rate. The value of this same dollar
at the end of two years is \((1 + i) + i(1 + i)\) or \((1 + i)^2\). Similarly, at the end of three years the value of one dollar is \((1 + i)^3\). This pattern continues for other years and mathematically the future value of one dollar is

\[
SPFW = (1 + i)^n
\]

where \(SPFW\) = single payment future worth ($/$)

\[i = \text{annual interest rate ($/$)}\]

\[n = \text{number of years (yr)}\]

The inverse of the single payment future worth is the present value of a dollar in some future year. This is commonly called the single payment present worth and it is defined by

\[
SPPW = \frac{1}{(1 + i)^n}
\]

where \(SPPW\) = single payment present worth ($/$)

\[i = \text{annual interest rate ($/$)}\]

\[n = \text{number of years (yr)}\]

Today's value of one dollar a year from now plus a dollar two years from now is \(1/(1 + i) + 1/(1 + i)^2\) per Equation 44. Similarly, today's value of three year end dollars is \(1/(1 + i) + 1/(1 + i)^2 + 1/(1 + i)^3\). This pattern continues for additional years and mathematically the present value of a series of one dollars in future years is

\[
USPW(i,n) = \frac{1}{(1 + i)} + \frac{1}{(1 + i)^2} + \frac{1}{(1 + i)^3} + \ldots + \frac{1}{(1 + i)^n}
\]

\[= \frac{(1 + i)^n - 1}{i(1 + i)^n}\]
where \( USPW(i,n) = \) uniform series present worth (\$/yr/$)
\( i = \) annual interest rate ($/$)
\( n = \) number of years (yr)

Since the proposed process is an unproven technology which will require a sizable capital investment, it is expected that the first application of this technology will occur in the governmental sector. For this reason, a 7 percent interest rate as proposed by the United States Department of Energy (DOE) will be utilized in this analysis [28]. This interest rate represents the real return on investment over and above inflation. Its use requires that the actual inflated cost of recurring expenses and fuel prices be estimated. DOE has also established guidelines for these two factors. For recurring costs like labor and materials, actual inflated costs are expected to rise at the average inflation rate [28]. Therefore, the present worth of recurring expenses is the product of current dollar cash flow and the uniform series present worth factor (Equation 45). Fuel prices however, are expected to rise at rates different than inflation depending on factors like fuel types, year of purchase, sector of economy, and region of the country [28]. Fuel types considered by this study are natural gas and electricity. The fuel escalation rates are established separately for each of three time periods. They are 1981 to 1985, 1985 to 1990, and 1990 to all outer years. This study covers all three time periods but the 1981 to 1985 period is the most significant because the escalation rates are highest during this period. The three types of residential, commercial and industrial customers
each have their own escalation rate. This study utilizes the industrial rate for the proposed process. Finally, escalation rates are provided for each area of the country to account for transportation cost. Region No. 4, which contains Florida was assumed. In all cases the future cost of fuel is predicted by

\[
FFC = TFC(1 + E1)^{Y1} (1 + E2)^{Y2} (1 + E3)^{Y3}
\]  
(46)

where \( FFC \) = future fuel cost ($/10E6 BTU)

\( TFC \) = today's fuel cost ($/10E6 BTU)

\( E1 \) = escalation rate 1981 to 1985, Table 8

\( E2 \) = escalation rate 1985 to 1990, Table 8

\( E3 \) = escalation rate 1990 to infinity, Table 8

\( Y1 \) = number of years in period 1981 to 1985 (yr)

\( Y2 \) = number of years in period 1985 to 1990 (yr)

\( Y3 \) = number of years in period 1990 to infinity (yr)

To obtain the present worth of all fuel it is necessary to predict fuel costs in each year by Equation 46, find the present worth of each of these fuel costs by Equation 44 and sum all the present worths.

Mathematically this relation is

\[
FSPWS = \sum_{YEAR=1982}^{DATE} \frac{FFC}{(1 + i)^{YEAR - 1981}}
\]  
(47)

where \( FSPWS \) = fuel series present worth sum ($-yr/10E6 BTU)

\( FFC \) = future fuel cost ($/10E6 BTU) Equation 46

\( i \) = annual interest rate ($/$)

\( YEAR \) = calculation period (yr)

\( DATE \) = last period in calculation (yr)

The results of Equation 47 are shown in Tables 9 and 10 for electricity and gas, respectively. Equations 45 and 47 both predict present worth which occurs one year in time before the first occurrence
<table>
<thead>
<tr>
<th>Fuel Type</th>
<th>1981 to 1985 (%)</th>
<th>1985 to 1990 (%)</th>
<th>1990 to Infinity (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electricity</td>
<td>5.28</td>
<td>1.85</td>
<td>.81</td>
</tr>
<tr>
<td>Natural gas</td>
<td>8.88</td>
<td>4.38</td>
<td>4.03</td>
</tr>
</tbody>
</table>

TABLE 9
ECONOMIC FACTORS FOR ELECTRICITY
(interest rate = 7% and TFC = $13.50/10E6 BTU)

<table>
<thead>
<tr>
<th>DATE</th>
<th>n</th>
<th>FFC (Eq. 46)</th>
<th>SPPW (Eq. 44)</th>
<th>FFCPW* (Eq. 47)</th>
<th>FSPWS (Eq. 47)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1982</td>
<td>1</td>
<td>14.213</td>
<td>0.935</td>
<td>13.289</td>
<td>13.289</td>
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<tr>
<td>1983</td>
<td>2</td>
<td>14.963</td>
<td>0.873</td>
<td>13.063</td>
<td>26.352</td>
</tr>
<tr>
<td>1984</td>
<td>3</td>
<td>15.753</td>
<td>0.816</td>
<td>12.854</td>
<td>39.205</td>
</tr>
<tr>
<td>1985</td>
<td>4</td>
<td>16.585</td>
<td>0.763</td>
<td>12.654</td>
<td>51.860</td>
</tr>
<tr>
<td>1986</td>
<td>5</td>
<td>16.892</td>
<td>0.713</td>
<td>12.044</td>
<td>63.904</td>
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<tr>
<td>1987</td>
<td>6</td>
<td>17.204</td>
<td>0.666</td>
<td>11.458</td>
<td>75.362</td>
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<tr>
<td>1988</td>
<td>7</td>
<td>17.523</td>
<td>0.623</td>
<td>10.917</td>
<td>86.279</td>
</tr>
<tr>
<td>1989</td>
<td>8</td>
<td>17.847</td>
<td>0.582</td>
<td>10.387</td>
<td>96.666</td>
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<tr>
<td>1990</td>
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<td>18.177</td>
<td>0.544</td>
<td>9.888</td>
<td>106.554</td>
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<td>1991</td>
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<td>18.324</td>
<td>0.508</td>
<td>9.309</td>
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<td>1992</td>
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<td>18.473</td>
<td>0.475</td>
<td>8.775</td>
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<tr>
<td>1993</td>
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<td>18.622</td>
<td>0.444</td>
<td>8.268</td>
<td>132.906</td>
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<tr>
<td>1994</td>
<td>13</td>
<td>18.773</td>
<td>0.415</td>
<td>7.791</td>
<td>140.697</td>
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<tr>
<td>1995</td>
<td>14</td>
<td>18.925</td>
<td>0.388</td>
<td>7.343</td>
<td>148.040</td>
</tr>
<tr>
<td>1996</td>
<td>15</td>
<td>19.078</td>
<td>0.362</td>
<td>6.906</td>
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<tr>
<td>1997</td>
<td>16</td>
<td>19.233</td>
<td>0.339</td>
<td>6.520</td>
<td>161.466</td>
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<tr>
<td>1998</td>
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<td>19.389</td>
<td>0.317</td>
<td>6.146</td>
<td>167.612</td>
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<tr>
<td>1999</td>
<td>18</td>
<td>19.546</td>
<td>0.296</td>
<td>5.786</td>
<td>173.398</td>
</tr>
<tr>
<td>2000</td>
<td>19</td>
<td>19.704</td>
<td>0.277</td>
<td>5.458</td>
<td>178.856</td>
</tr>
<tr>
<td>2001</td>
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<td>19.864</td>
<td>0.258</td>
<td>5.125</td>
<td>183.981</td>
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<tr>
<td>2002</td>
<td>21</td>
<td>20.025</td>
<td>0.242</td>
<td>4.846</td>
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<tr>
<td>2003</td>
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<td>20.187</td>
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<td>193.389</td>
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<tr>
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<td>20.350</td>
<td>0.211</td>
<td>4.294</td>
<td>197.683</td>
</tr>
<tr>
<td>2005</td>
<td>24</td>
<td>20.515</td>
<td>0.197</td>
<td>4.041</td>
<td>201.724</td>
</tr>
<tr>
<td>2006</td>
<td>25</td>
<td>20.681</td>
<td>0.184</td>
<td>3.805</td>
<td>205.529</td>
</tr>
<tr>
<td>2007</td>
<td>26</td>
<td>20.849</td>
<td>0.172</td>
<td>3.586</td>
<td>209.115</td>
</tr>
<tr>
<td>2008</td>
<td>27</td>
<td>21.018</td>
<td>0.161</td>
<td>3.394</td>
<td>212.499</td>
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<tr>
<td>2009</td>
<td>28</td>
<td>21.188</td>
<td>0.150</td>
<td>3.178</td>
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<tr>
<td>2010</td>
<td>29</td>
<td>21.360</td>
<td>0.141</td>
<td>3.012</td>
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<tr>
<td>2011</td>
<td>30</td>
<td>21.533</td>
<td>0.131</td>
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<tr>
<td>2012</td>
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<td>21.707</td>
<td>0.123</td>
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<td>2013</td>
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<td>21.883</td>
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<td>226.697</td>
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<tr>
<td>2014</td>
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<td>22.060</td>
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<td>2015</td>
<td>34</td>
<td>22.239</td>
<td>0.100</td>
<td>2.224</td>
<td>231.281</td>
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<tr>
<td>2016</td>
<td>35</td>
<td>22.419</td>
<td>0.940</td>
<td>2.107</td>
<td>233.388</td>
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TABLE 9 - Continued

<table>
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<tr>
<th>DATE</th>
<th>n</th>
<th>FFC (Eq. 46)</th>
<th>SPPW (Eq. 44)</th>
<th>FFCPW* (Eq. 47)</th>
<th>FSPWS</th>
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</thead>
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<tr>
<td>2017</td>
<td>36</td>
<td>22.601</td>
<td>0.088</td>
<td>1.989</td>
<td>235.377</td>
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<tr>
<td>2018</td>
<td>37</td>
<td>22.784</td>
<td>0.082</td>
<td>1.868</td>
<td>237.245</td>
</tr>
<tr>
<td>2019</td>
<td>38</td>
<td>22.968</td>
<td>0.076</td>
<td>1.746</td>
<td>238.991</td>
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<td>2020</td>
<td>39</td>
<td>23.154</td>
<td>0.071</td>
<td>1.644</td>
<td>240.635</td>
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<tr>
<td>2021</td>
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<td>23.342</td>
<td>0.067</td>
<td>1.564</td>
<td>242.199</td>
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<tr>
<td>2023</td>
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<td>23.721</td>
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<td>2024</td>
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* FFCPW = FFC * SPPW
TABLE 10

ECONOMIC FACTORS FOR NATURAL GAS
(interest rate = 7% and TFC = $6.00/10E6 BTU)

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1986 5 8.802 0.713 6.276 31.348
1987 6 9.187 0.666 6.119 37.467
1988 7 9.589 0.623 5.974 43.441
1989 8 10.010 0.582 5.826 49.267
1990 9 10.448 0.544 5.684 54.951
1991 10 10.869 0.500 5.521 60.472
1992 11 11.307 0.475 5.371 65.843
1993 12 11.763 0.444 5.223 71.066
1994 13 12.237 0.415 5.078 76.144
1995 14 12.730 0.388 4.939 81.083
1996 15 13.243 0.362 4.794 85.877
1997 16 13.777 0.339 4.670 90.547
1998 17 14.332 0.317 4.543 95.090
1999 18 14.909 0.296 4.413 99.503
2000 19 15.510 0.277 4.296 103.799
2001 20 16.135 0.258 4.163 107.962
2002 21 16.785 0.242 4.062 112.024
2003 22 17.462 0.226 3.946 115.970
2004 23 18.166 0.211 3.833 119.803
2005 24 18.898 0.197 3.723 123.526
2006 25 19.659 0.184 3.617 127.143
2007 26 20.452 0.172 3.518 130.661
2008 27 21.276 0.161 3.425 134.086
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Table 10 - Continued

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* FFCPW = FFC * SPPW
in the series. Therefore to combine series payments with lump sum payments like capital cost, this analysis assumes that all recurring costs and fuel costs occur on the last day of the year. In addition, it is assumed that the entire construction cost occurs on the first day of plant operation. These two assumptions simplify the economic analysis and Equation 48 can be solved for the payoff period. Payoff period is the number of years which equates the present worth of gas sales revenues to the present worth sum of electricity, construction and recurring costs or

\[
X \times FSPWS_{\text{gas}} = E \times FSPWS_{\text{elec}} + \text{Const.} + \text{ARC} \times USPW
\]  

(48)

where

- \(X\) = plant gas output per year (BTU/yr)
- \(FSPWS_{\text{gas}}\) = fuel series present worth sum for gas ($-yr/10E6 BTU)
- \(E\) = plant electricity consumption per year (BTU/yr)
- \(FSPWS_{\text{elec}}\) = fuel series present worth sum for electricity ($-yr/10E6 BTU)
- \(\text{Const.}\) = construction cost ($)
- \(\text{ARC}\) = annual recurring costs ($/yr)
- \(USPW\) = uniform series present worth ($-yr/$)
IV. RESULTS

This chapter presents the analytical results for 0.1, 0.5 and 1.0(10E12) BTU per year methane production systems. Three sizes were selected because they represent capacities which approximate 10, 50 and 100 percent consumption levels typical of a Central Florida gas utility [2]. In addition, competitively priced gas clean-up equipment is not readily available for gas flow rates smaller than 1.5(10E6) SCF per day which is near the size required for the 0.5(10E12) BTU per year system. Most of the gas clean-up technology has been developed for coal to gas systems which are orders of magnitude larger than these systems. Three sizes of conversion facilities were analyzed so that trends (the effect of scale) in energy consumption and cost could be analyzed. It is typical for most systems to change their efficiency as their size or capacity changes. The proposed methane production process is totally unproven in the commercial sector and hence, substantial amounts of basic research will be necessary to introduce this technology. Identification of trends can provide the insight which will allow the most effective utilization of research dollars and man power. This should promote faster introduction of the technology as well as lower overall gas cost.
In the discussion which follows the data will be divided into the two sub-systems of conversion facility and farm. This will allow the comparison of individual system data from other studies with the results of this work. Specifically, certain regions of the country may have local waste sources (e.g. agricultural packing house wastes) which might be available at no or limited costs. With the results separated into farm costs and conversion cost others will be able to examine the economics of their alternative feedstocks with minimum effort.

The results will be presented in four subsections. The first contains the major component sizes which provides insight into the system's physical size and also forms the basis of the capital cost estimates. Subsection two details the energy consumption of the major components and establishes the net energy production. The third subsection determines the capital cost, operating/maintenance costs and payoff period. Finally, the last subsection discusses the implications of the results and proposes a prioritized list of research goals. This list will allow research to be focused on key areas and the result should be faster development of a renewable energy source at the lowest possible cost.

**Equipment Size Summary**

Piping was sized using the allowable velocities developed from Equation 6. Table 11 shows the recommended size and length of the major stream flows identified in Figure 2 and Table 2. The results
### TABLE 11

**PIPE SIZES**

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<th>Stream Flow No.</th>
<th>Description</th>
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<th>Dia- Meter Length</th>
<th>Dia- Meter Length</th>
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<td>0.5 (10E12) BTU/yr</td>
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</table>

* This is a solids stream.
indicate that generally 0.5 to 14 inch piping will be necessary. This range of pipe sizes is commonly used in process plants. Hence, the piping portion of the construction effort should present no new technological problems. The exception to the above is the low pressure gas collection piping from the methane phase digester (MPD) to the compressor. Here pipe sizes become quite large (10 to 28 inch) and although this size piping is conventional equipment, its cost can be large. To minimize this cost component, the engineering design effort should give considerable thought to the geometric relationship of the MPD and compressor. Of the assumptions discussed in Chapter III, only average pipe size is sensitive to scale. Average pipe size was used to calculate allowable velocity. A value of 4 inches was assumed and inspection of Table 11 indicates that this assumption was reasonable for all flows of the 0.1(10E12) BTU per year facility except the low pressure gas lines. Average pipe size is related to allowable velocity through the friction factor in Equation 6. For 4 inch pipe $f = 0.016$ while for 20 inch pipe $f = 0.012$. This difference would have the effect of increasing velocity 33 percent and increasing frictional pressure drop 77 percent. Since the original frictional pressure drop was an extremely low value (1.0 psi per 100 feet), the total increase in pressure drop is insignificant when compared to the uncertainty of the whole analysis. For this reason the piping sizes indicated in Table 11 are considered satisfactory as presented.

Vessels except for the leaching-bed reactors were sized to provide 1.5 days hydraulic retention time. The leaching-bed reactors
were sized using Equation 8. Table 12 shows the calculated vessel sizes. The results show that the leaching-beds are quite massive and hence, considerable land area will be required for the conversion facility. Of the assumptions discussed in Chapter III both tank height (limited to 30 feet) and the 1.5 compaction factor are sensitive to scale. Data on the compaction of biomass as it digests is unavailable because the proposed process is completely unproven. It is expected that research in progress will show the compaction factor to be somewhat larger. If this happens the leaching-bed could be sized proportionately smaller. The 30 feet tank height is necessary to minimize pumping head but for the 1.0(10E12) BTU per year system it seems more realistic to specify multiples of small tanks rather than one large tank. Table 13 shows recommended tank sizes which utilize multiple units. Since the average tank size exceeds 30^3 feet^3, little cost differential is expected. Using multiple tanks has the additional benefit of adding operational flexibility to the system.

Pump sizes were determined from the data of Table 4 and Equation 14. Pump capacities for the larger systems are multiples of the values in Table 4. Since tank height is constrained to 30 feet, the pump heads shown in Table 4 apply to all system sizes. Pump power requirements were estimated by Equation 14. The driver sizes were established by rounding the power requirement up to the next larger commercial electric motor size. Table 14 details the recommended pump and driver sizes. The results show that except for pump No. 9 (biomass supply pump), all pumps are less than 25 horsepower which makes
### TABLE 12

**CALCULATED VESSELS SIZES**

<table>
<thead>
<tr>
<th>Tank No.</th>
<th>Description</th>
<th>System Size = 0.1 (10E12 BTU/yr)</th>
<th>System Size = 0.5 (10E12 BTU/yr)</th>
<th>System Size = 1.0 (10E12 BTU/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Volume (10^3 ft³)</td>
<td>Length (ft)</td>
<td>Volume (10^3 ft³)</td>
</tr>
<tr>
<td>1.</td>
<td>Acid phase digester</td>
<td>1,323</td>
<td>210</td>
<td>6,615</td>
</tr>
<tr>
<td>2.</td>
<td>Acid storage tank</td>
<td>32</td>
<td>33</td>
<td>161</td>
</tr>
<tr>
<td>3.</td>
<td>Methane phase digester</td>
<td>32</td>
<td>33</td>
<td>161</td>
</tr>
<tr>
<td>4.</td>
<td>Recycle water tank</td>
<td>32</td>
<td>33</td>
<td>161</td>
</tr>
<tr>
<td>5.</td>
<td>Liquid storage tank</td>
<td>26</td>
<td>30</td>
<td>132</td>
</tr>
<tr>
<td>6.</td>
<td>Lime slurry tank</td>
<td>1</td>
<td>5</td>
<td>4</td>
</tr>
</tbody>
</table>

**NOTE:** Vessels are assumed 30 feet high; Length = (Volume/30)^0.5
<table>
<thead>
<tr>
<th>Tank No.</th>
<th>Description</th>
<th>System Size =</th>
<th>0.1 (10E12) BTU/yr</th>
<th>0.5 (10E12) BTU/yr</th>
<th>1.0 (10E12) BTU/yr</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Base Quantity (ft)</td>
<td>Base Quantity (ft)</td>
<td>Base Quantity (ft)</td>
</tr>
<tr>
<td>1.</td>
<td>Acid phase digester</td>
<td></td>
<td>1 210</td>
<td>1 470</td>
<td>2 470</td>
</tr>
<tr>
<td>2.</td>
<td>Acid storage tank</td>
<td></td>
<td>1 33</td>
<td>1 73</td>
<td>2 73</td>
</tr>
<tr>
<td>3.</td>
<td>Methane phase digester</td>
<td></td>
<td>1 33</td>
<td>1 73</td>
<td>2 73</td>
</tr>
<tr>
<td>4.</td>
<td>Recycle water tank</td>
<td></td>
<td>1 33</td>
<td>1 73</td>
<td>2 73</td>
</tr>
<tr>
<td>5.</td>
<td>Liquid storage tank</td>
<td></td>
<td>1 30</td>
<td>1 66</td>
<td>2 66</td>
</tr>
<tr>
<td>6.</td>
<td>Lime slurry tank</td>
<td></td>
<td>1 5</td>
<td>1 11</td>
<td>2 11</td>
</tr>
</tbody>
</table>
### TABLE 14

**PUMP SIZES**

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Flow*  (GPM)</th>
<th>Head* (ft)</th>
<th>Power** (hp)</th>
<th>Driver*** (hp)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>198</td>
<td>24</td>
<td>2.3</td>
<td>3</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>111</td>
<td>24</td>
<td>1.3</td>
<td>2</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>110</td>
<td>24</td>
<td>1.3</td>
<td>2</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>12</td>
<td>24</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>91</td>
<td>24</td>
<td>1.0</td>
<td>1</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>3</td>
<td>39</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>3</td>
<td>24</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>3</td>
<td>24</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>204</td>
<td>1,880</td>
<td>182.1</td>
<td>200</td>
</tr>
</tbody>
</table>

---

**Size X = 0.5(10E12) BTU/yr**

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Flow*  (GPM)</th>
<th>Head* (ft)</th>
<th>Power** (hp)</th>
<th>Driver*** (hp)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>990</td>
<td>24</td>
<td>11</td>
<td>15</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>556</td>
<td>24</td>
<td>6.3</td>
<td>7.5</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>555</td>
<td>24</td>
<td>6.3</td>
<td>7.5</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>60</td>
<td>24</td>
<td>0.7</td>
<td>1.0</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>455</td>
<td>24</td>
<td>5.2</td>
<td>7.5</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>15</td>
<td>39</td>
<td>0.3</td>
<td>0.5</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>15</td>
<td>24</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>15</td>
<td>24</td>
<td>0.1</td>
<td>0.5</td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>1,020</td>
<td>1,880</td>
<td>910.4</td>
<td>1,000</td>
</tr>
</tbody>
</table>

---

**Size X = 1.0(10E12) BTU/yr**

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Flow*  (GPM)</th>
<th>Head* (ft)</th>
<th>Power** (hp)</th>
<th>Driver*** (hp)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>1,980</td>
<td>24</td>
<td>22.6</td>
<td>15</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>1,111</td>
<td>24</td>
<td>12.6</td>
<td>15</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>1,110</td>
<td>24</td>
<td>12.6</td>
<td>15</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>120</td>
<td>24</td>
<td>1.4</td>
<td>2.0</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>910</td>
<td>24</td>
<td>10.4</td>
<td>15</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>30</td>
<td>39</td>
<td>.5</td>
<td>1.0</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>30</td>
<td>24</td>
<td>.3</td>
<td>0.5</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>20</td>
<td>24</td>
<td>.3</td>
<td>0.5</td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>2,040</td>
<td>1,880</td>
<td>1,820.8</td>
<td>2,000</td>
</tr>
</tbody>
</table>

* Per Table 4  
** Per Equation 14  
*** Next larger standard electric motor size
them small when compared to many processing plants. The biomass supply pump on the other hand has a power requirement of 182 to 1,821 horsepower. These are significant size pumps and efforts to optimize their efficiency and cost should be undertaken at the time of engineering design. Of the assumptions discussed in Chapter III, only the electric motor efficiency for the biomass supply pump is significantly affected by plant size. A value of .82 was used to prepare Table 14 but for the larger system sizes, electric motor efficiency should be closer to .94. This discrepancy would result in a 12.5 percent smaller motor size. Since the estimated power requirement is a linear function of pipe length and since the 5,300 feet effective length assumption could easily be low by 12.5 percent, the predicted horsepower will be retained as the preliminary design basis.

Conveyors were sized using Equation 16, the required weight flow rate of biomass (Table 2, column 1, row 1), the chopped density of water hyacinth (Table 3), and a 500 feet per minute belt velocity. From Table 2 the required biomass handling rates are 51, 254 and 508 tons per hour for the 0.1, 0.5 and 1.0(10E12) BTU per year systems, respectively. This is based on 12 hours operation per day. Table 15 shows capacities for various width belt conveyors. Based on Table 15 the recommended conveyor sizes are as follows:

<table>
<thead>
<tr>
<th>System Size (10E12 BTU/yr)</th>
<th>Belt Conveyor Width (in)</th>
</tr>
</thead>
<tbody>
<tr>
<td>X = 0.1</td>
<td>18</td>
</tr>
<tr>
<td>X = 0.5</td>
<td>24</td>
</tr>
<tr>
<td>X = 1.0</td>
<td>36</td>
</tr>
</tbody>
</table>
### TABLE 15

**BELT CONVEYOR CAPACITIES**

<table>
<thead>
<tr>
<th>Belt Width (in)</th>
<th>Capacity (ton/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>18</td>
<td>135</td>
</tr>
<tr>
<td>24</td>
<td>250</td>
</tr>
<tr>
<td>30</td>
<td>400</td>
</tr>
<tr>
<td>36</td>
<td>585</td>
</tr>
<tr>
<td>42</td>
<td>825</td>
</tr>
</tbody>
</table>

**NOTES:** For chopped water hyacinth with 50 ft³/lb density and 8.33 ft/sec belt speed.
The driver sizes were estimated from Equation 21 which requires knowledge of the conveyor's geometry and capacity. Conveyor geometry is composed of the conveyor's length and its change in elevation (height). This analysis assumes the conveyors in all three size systems will be 10 feet high. Since the exact conveyor arrangement is unknown, 10 feet was selected to provide clearance between the conveyor discharge and the biomass supply pump's sump. The conveyor length is the same as the collection boom's circumference. Figure 6 shows this relationship. The collection boom configuration is defined by Equation 25 and the required weight flow rate (Table 2, column 1, row 1). Table 16 shows the specifics of the fan area including collector circumference. From these assumptions, Table 17 was developed from equation 21. It shows the contribution of each power component as well as the total power requirement and the driver size for each conveyor. Table 17 also shows that the parasitic load to keep an empty belt moving is a significantly larger percentage of the total input power for the 0.1(10E12) BTU per year system. Figure 8 shows this trend graphically. Interestingly this trend is not of concern since total input power for this size is only 15.1 horsepower. Of the assumptions discussed in Chapter III none appear sensitive to scale.

Winches were sized by applying Equation 29 while neglecting friction drag as outlined in Chapter III. To apply Equation 29, it is necessary to determine the fan size, the maximum towing velocity, relative air velocity and relative water velocity. Most of these factors are functions of plant size and their determination is
### Table 16

**FARM AREA CONFIGURATION**

<table>
<thead>
<tr>
<th>System Size (BTU/yr)</th>
<th>Feedstock Requirement (dry ton/yr)</th>
<th>Harvest Area (Eq. 24 acre/day)</th>
<th>Boom Diameter (Eq. 25 ft)</th>
<th>( \frac{1}{2} ) Circumference (ft)</th>
<th>Farm Area (acre)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1(10E12)</td>
<td>11,133</td>
<td>3.7</td>
<td>637</td>
<td>1,001</td>
<td>291</td>
</tr>
<tr>
<td>0.5(10E12)</td>
<td>55,663</td>
<td>18.3</td>
<td>1,425</td>
<td>2,238</td>
<td>1,455</td>
</tr>
<tr>
<td>1.0(10E12)</td>
<td>111,325</td>
<td>36.6</td>
<td>2,016</td>
<td>3,167</td>
<td>2,910</td>
</tr>
</tbody>
</table>

* (Table 2, column 1, row 1) times 365 days.

### Table 17

**CONVEYOR POWER REQUIREMENTS**

<table>
<thead>
<tr>
<th>System Size (BTU/yr)</th>
<th>Belt Length (ft)</th>
<th>Belt Speed (ft/min)</th>
<th>Belt Capacity (ton/hr) (Eq. 16)</th>
<th>Input Power Empty (hp)</th>
<th>Horiz. (hp)</th>
<th>Vert. (hp)</th>
<th>Total (hp) (Eq. 21)</th>
<th>Driver* Size (hp)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1(10E12)</td>
<td>1,001</td>
<td>500</td>
<td>51</td>
<td>12.2</td>
<td>2.3</td>
<td>0.6</td>
<td>15.1</td>
<td>15</td>
</tr>
<tr>
<td>0.5(10E12)</td>
<td>2,238</td>
<td>500</td>
<td>254</td>
<td>35.5</td>
<td>23.8</td>
<td>3.1</td>
<td>62.4</td>
<td>75</td>
</tr>
<tr>
<td>1.0(10E12)</td>
<td>3,167</td>
<td>500</td>
<td>508</td>
<td>90.8</td>
<td>54.3</td>
<td>6.2</td>
<td>151.3</td>
<td>200</td>
</tr>
</tbody>
</table>

* Next larger standard electric motor size.
<table>
<thead>
<tr>
<th>WORK TYPE</th>
<th>SYSTEM SIZE</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.1(10E12) BTU/yr</td>
</tr>
<tr>
<td>HORIZONTAL</td>
<td>4.0%</td>
</tr>
<tr>
<td>VERTICAL</td>
<td>15.2%</td>
</tr>
<tr>
<td>EMPTY</td>
<td>80.8%</td>
</tr>
</tbody>
</table>

Figure 8. Comparison of Conveyor Power Requirements
discussed in Chapter III. Table 18 presents both the above factor as well as input power by component and driver size for each system size. There are a total of four winches required to operate the collection system but only two will operate at any given time. Driver size was determined by dividing total input power in half and rounding down to the nearest standard electric motor size. Rounding down is reasonable since the major winch power component is air form drag. This drag results from selecting a 94 feet per second wind speed. This value is the highest wind speed ever recorded in Orlando, Florida whereas 69 feet per second is the average maximum wind speed in a normal year [51]. On the rare occasions when wind speed is excessive harvesting will probably be discontinued anyway. Table 18 also shows that water form drag becomes a significant factor with increasing towing speeds. This results from longer pulling distances which must be traversed in the same 3 hour towing period. Of the assumptions discussed in Chapter III only neglecting friction drag has the potential to be sensitive to scale. A rough calculation indicates that frictional drag for the 1.0(10E12) BTU per year system (the worst case) would increase power requirements by about 70 percent. If the normal average wind speed (69 feet per second) had been used in winch sizing, roughly the same size winches would have been recommended. Therefore, neglecting friction drag is reasonable when sizing winches but energy consumption calculations should consider this force.

Gas-cleaning and compression equipment were not sized in detail as their requirements are established by the mass balance Table 2.
**TABLE 18**

**WINCH POWER REQUIREMENTS**

<table>
<thead>
<tr>
<th>System Size =</th>
<th>0.1</th>
<th>0.5</th>
<th>1.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>(10E12)</td>
<td>(10E12)</td>
<td>(10E12)</td>
<td></td>
</tr>
<tr>
<td>BTU/yr</td>
<td>BTU/yr</td>
<td>BTU/yr</td>
<td></td>
</tr>
</tbody>
</table>

**Description**

<table>
<thead>
<tr>
<th>Description</th>
<th>0.1</th>
<th>0.5</th>
<th>1.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Travel distance maximum (ft)*</td>
<td>5,035</td>
<td>11,259</td>
<td>15,922</td>
</tr>
<tr>
<td>Maximum velocity to shore (ft/sec)**</td>
<td>0.47</td>
<td>1.04</td>
<td>1.47</td>
</tr>
<tr>
<td>Relative design velocity</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water (ft/sec)</td>
<td>1.0</td>
<td>1.5</td>
<td>2.0</td>
</tr>
<tr>
<td>Air (ft/sec)</td>
<td>94</td>
<td>94</td>
<td>94</td>
</tr>
<tr>
<td>Input power</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water (hp)</td>
<td>0.951</td>
<td>10.54</td>
<td>37.53</td>
</tr>
<tr>
<td>Air (hp)</td>
<td>16.99</td>
<td>84.07</td>
<td>168.01</td>
</tr>
<tr>
<td>Total (hp) Eq. 29</td>
<td>17.94</td>
<td>94.61</td>
<td>205.54</td>
</tr>
<tr>
<td>Driver size (hp)***</td>
<td>10</td>
<td>50</td>
<td>50</td>
</tr>
</tbody>
</table>

* Length of triangle base.
** To travel maximum distance in 3 hours.
*** Total power is split between two winches.
The driver size for compressors was established using Equation 34. Table 19 summarizes the gas preparation equipment data. The table shows that unlike most other plant equipment which have power requirements less than 50 horsepower, the compressors have significant power requirements. Efforts to optimize compressor efficiency and cost would probably yield significant capital and operating cost savings. None of the assumptions discussed in Chapter III appear to be sensitive to scale.

**Energy Consumption**

Net energy production is necessary for the proposed process to be of value. Chapter III previously showed that net energy production occurs if Equation 42 is satisfied. This subsection develops the data for Equation 36 by discussing the work components individually.

The yearly energy consumed by pumps is the product of power requirement and operating time as outlined by Equation 37. Since the conversion facility's service factor is 0.88, operating time is assumed to be 90 percent of a year's 8,760 hours. This is necessary to account for the energy consumed during start-up and shut-down periods. Table 20 details the yearly energy consumption of each pump. It should be noted that the biomass supply pump accounts for over 90 percent of all pump energy consumption. The yearly energy consumption of conveyors is shown in Table 21 as predicted by Equation 38. Gas compressor energy consumption is outlined in Table 22 as calculated
<table>
<thead>
<tr>
<th>System Size (BTU/yr)</th>
<th>Compressor* Capacity (SCF/day)</th>
<th>Weight Flow** (lb mole/sec)</th>
<th>Input Power Eq. 34 (hp)</th>
<th>Driver*** Size (hp)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1(10E12)</td>
<td>365,900</td>
<td>0.0088</td>
<td>90.3</td>
<td>100</td>
</tr>
<tr>
<td>0.5(10E12)</td>
<td>1,829,500</td>
<td>0.0438</td>
<td>449</td>
<td>500</td>
</tr>
<tr>
<td>1.0(10E12)</td>
<td>3,659,000</td>
<td>0.0878</td>
<td>900</td>
<td>1,000</td>
</tr>
</tbody>
</table>

* Per Table 2, column 1, row 1.
** Compressor capacity (from Table 2) times average gas density divided by average mole weight divided by 86,400 sec/day. Average gas density for methane and carbon dioxide mixture is .0526 lb/ft³ while average mole weight is 25.4 lb/lb mole.
*** Next size standard electric motor size.
<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Operating Time (hr)</th>
<th>Pp (hp)</th>
<th>Ep (hp-hr/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>7,884</td>
<td>2.3</td>
<td>18,133</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>7,884</td>
<td>1.3</td>
<td>10,249</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>7,884</td>
<td>1.3</td>
<td>10,249</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>7,884</td>
<td>1.0</td>
<td>7,884</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td></td>
<td>Subtotal conversion facility</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>3,942</td>
<td>182.1</td>
<td>717,838</td>
</tr>
<tr>
<td></td>
<td>Subtotal fan</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Total system</td>
<td></td>
<td></td>
<td>767,507</td>
</tr>
</tbody>
</table>

Size X = 0.1(10E12) BTU/yr

Size X = 0.5(10E12) BTU/yr

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Operating Time (hr)</th>
<th>Pp (hp)</th>
<th>Ep (hp-hr/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>7,884</td>
<td>11.0</td>
<td>86,724</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>7,884</td>
<td>6.3</td>
<td>49,669</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>7,884</td>
<td>6.3</td>
<td>49,669</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>7,884</td>
<td>0.7</td>
<td>5,519</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>7,884</td>
<td>5.2</td>
<td>40,997</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>7,884</td>
<td>0.3</td>
<td>2,365</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>7,884</td>
<td>0.1</td>
<td>788</td>
</tr>
<tr>
<td></td>
<td>Subtotal conversion facility</td>
<td></td>
<td></td>
<td>236,520</td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>3,942</td>
<td>910.4</td>
<td>3,588,797</td>
</tr>
<tr>
<td></td>
<td>Subtotal fan</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Total system</td>
<td></td>
<td></td>
<td>3,825,317</td>
</tr>
</tbody>
</table>
### TABLE 20 - Continued

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Description</th>
<th>Operating Time (hr)</th>
<th>Pp (hp)</th>
<th>Ep (hp-hr/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P-1</td>
<td>Acid phase digester discharge</td>
<td>7,884</td>
<td>22.6</td>
<td>178,178</td>
</tr>
<tr>
<td>P-2</td>
<td>Acid storage tank discharge</td>
<td>7,884</td>
<td>12.6</td>
<td>99,338</td>
</tr>
<tr>
<td>P-3</td>
<td>Recycle storage tank discharge</td>
<td>7,884</td>
<td>12.6</td>
<td>99,338</td>
</tr>
<tr>
<td>P-4</td>
<td>Acid surge tank solids</td>
<td>7,884</td>
<td>1.4</td>
<td>11,038</td>
</tr>
<tr>
<td>P-5</td>
<td>Liquid storage tank discharge</td>
<td>7,884</td>
<td>10.4</td>
<td>81,994</td>
</tr>
<tr>
<td>P-6</td>
<td>Supply water</td>
<td>7,884</td>
<td>0.5</td>
<td>3,942</td>
</tr>
<tr>
<td>P-7</td>
<td>Lime slaker discharge</td>
<td>7,884</td>
<td>0.3</td>
<td>2,365</td>
</tr>
<tr>
<td>P-8</td>
<td>Lime slurry tank discharge</td>
<td>7,884</td>
<td>0.3</td>
<td>2,365</td>
</tr>
<tr>
<td></td>
<td>Subtotal conversion facility</td>
<td></td>
<td></td>
<td>478,559</td>
</tr>
<tr>
<td>P-9</td>
<td>Biomass supply pump</td>
<td>3,942</td>
<td>1,820.8</td>
<td>7,177,594</td>
</tr>
<tr>
<td></td>
<td>Subtotal fann</td>
<td></td>
<td></td>
<td>7,177,594</td>
</tr>
<tr>
<td></td>
<td><strong>Total system</strong></td>
<td></td>
<td></td>
<td><strong>7,656,153</strong></td>
</tr>
</tbody>
</table>

**NOTES:**
- Pp = Pump input power from Table 14 (hp).
- Ep = Yearly pump energy (hp-hr/yr) per Equation 37.

Size X = 1.0(10E12) BTU/yr
### TABLE 21

**CONVEYOR YEARLY ENERGY CONSUMPTION**

<table>
<thead>
<tr>
<th>System Size =</th>
<th>0.1 (10E12) BTU/yr</th>
<th>0.5 (10E12) BTU/yr</th>
<th>1.0 (10E12) BTU/yr</th>
</tr>
</thead>
<tbody>
<tr>
<td>Description</td>
<td>Operating Time (hr)</td>
<td>Pc (hp)</td>
<td>Ec (hp-hr/yr)</td>
</tr>
<tr>
<td>Shore conveyor at fam</td>
<td>3,942</td>
<td>15.1</td>
<td>59,524</td>
</tr>
<tr>
<td>Residual solids conveyor at conversion facility</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total system</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**NOTES:**
- Pc = Conveyor input power from Table 17 (hp).
- Ec = Yearly conveyor energy (hp-hr/yr) per Equation 38.
- * 11 percent of shore conveyor's consumption.

### TABLE 22

**COMPRESSOR YEARLY ENERGY CONSUMPTION**

<table>
<thead>
<tr>
<th>System size =</th>
<th>0.1 (10E12) BTU/yr</th>
<th>0.5 (10E12) BTU/yr</th>
<th>1.0 (10E12) BTU/yr</th>
</tr>
</thead>
<tbody>
<tr>
<td>Description</td>
<td>Operating Time (hr)</td>
<td>Px (hp)</td>
<td>Ex (hp-hr/yr)</td>
</tr>
<tr>
<td>Gas compressor 15 to 250 psia</td>
<td>7,884</td>
<td>90.3</td>
<td>711,925</td>
</tr>
</tbody>
</table>

**NOTES:**
- Px = Compressor input power from Table 19 (hp).
- Ex = Yearly compressor energy (hp-hr/yr) per Equation 41.
from Equation 41. Yearly energy consumption for grinding is shown in Table 23 and was calculated from Equation 40. Finally, Table 24 outlines the yearly energy consumption of winches as predicted by Equation 39. Recall that the maximum power requirement of winches was dominated by wind drag. However, the energy consumption is most influenced by water drag. This results from the large difference between peak wind velocity and average wind velocity.

A summary of energy consumption is shown in Table 25. Inspection of this table reveals that the percentage contribution of each consumer remains approximately the same. The average percent distribution of energy consumption is shown in Figure 9. From this it is clear that the compressor and the biomass supply pump are the major consumers. Together they account for about 90 percent of the energy consumed by the process. Generally there are no significant effects of scale on total process energy consumption. In other words, the unit energy consumption per cubic foot of gas is relatively constant. Figure 10 plots normalized component energy consumption (energy of size X divided by energy of size 1.0) and shows the effects of scale on component energy consumption. Individually consumers like the winches and the conveyors exhibit a trend towards increasing unit energy consumption with system size because of increasing transportation distances. However, these components represent such a small portion (less than 4 percent) of the total consumption that this trend does not develop in the total process consumption. Note that
### TABLE 23

**GRINDER YEARLY ENERGY CONSUMPTION**

<table>
<thead>
<tr>
<th>System Size (BTU/yr)</th>
<th>DTY (dry ton/yr)</th>
<th>Eg (hp-hr/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1(10E12)</td>
<td>9,690</td>
<td>96,900</td>
</tr>
<tr>
<td>0.5(10E12)</td>
<td>48,450</td>
<td>484,500</td>
</tr>
<tr>
<td>1.0(10E12)</td>
<td>96,900</td>
<td>969,900</td>
</tr>
</tbody>
</table>

**NOTES:**
- DTY = Yearly feedstock requirement (dry ton/yr) per Equation 22.
- Eg = Yearly energy consumption of grinder (hp-hr/yr) per Equation 46.

### TABLE 24

**WINCH YEARLY ENERGY CONSUMPTION**

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Water (ft/sec)</td>
<td>DF1 (lb)</td>
<td>DF2 (lb)</td>
<td>DF3 (lb)</td>
<td>DF4 (lb)</td>
<td>DFT (lb)</td>
<td>l (ft)</td>
<td>Ew (hp-hr/yr)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.1(10E12)</td>
<td>.47</td>
<td>2.0</td>
<td>6</td>
<td>2</td>
<td>161</td>
<td>68</td>
<td>237</td>
<td>5,035</td>
<td>474</td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.5(10E12)</td>
<td>1.04</td>
<td>2.5</td>
<td>21</td>
<td>18</td>
<td>1,759</td>
<td>1,673</td>
<td>3,471</td>
<td>11,259</td>
<td>15,531</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.0(10E12)</td>
<td>1.47</td>
<td>3.0</td>
<td>42</td>
<td>50</td>
<td>4,973</td>
<td>6,682</td>
<td>11,747</td>
<td>15,922</td>
<td>74,330</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**NOTES:**
- DF1, DF2, DF3, & DF4 = Drag force per Equation 23.
- DFT = Total drag force (lb) per Equation 25.
- l = Length of triangle base (ft) per Table 18.
- Ew = Yearly energy consumption of winches (hp-hr/yr) per Equation 39 with towing speed toward shore per Table 18.
# TABLE 25

## TOTAL YEARLY ENERGY CONSUMPTION

<table>
<thead>
<tr>
<th>System Size =</th>
<th>0.1 (10E12) BTU/yr</th>
<th>0.5 (10E12) BTU/yr</th>
<th>1.0 (10E12) BTU/yr</th>
</tr>
</thead>
<tbody>
<tr>
<td>Description</td>
<td>(hp-hr/yr)</td>
<td>(hp-hr/yr)</td>
<td>(hp-hr/yr)</td>
</tr>
<tr>
<td>Pump Nos. 1 to 8</td>
<td>49,669</td>
<td>236,520</td>
<td>478,559</td>
</tr>
<tr>
<td>Conveyor (residual solids)</td>
<td>6,548</td>
<td>27,058</td>
<td>65,607</td>
</tr>
<tr>
<td>Compressor</td>
<td>711,925</td>
<td>3,539,916</td>
<td>7,805,160</td>
</tr>
<tr>
<td>Subtotal</td>
<td>768,142</td>
<td>3,803,494</td>
<td>8,349,326</td>
</tr>
<tr>
<td>Pump No. 9</td>
<td>717,838</td>
<td>3,588,797</td>
<td>7,177,594</td>
</tr>
<tr>
<td>Conveyor (shore)</td>
<td>59,524</td>
<td>245,981</td>
<td>596,425</td>
</tr>
<tr>
<td>Grinders</td>
<td>96,900</td>
<td>484,500</td>
<td>969,000</td>
</tr>
<tr>
<td>Winch</td>
<td>474</td>
<td>15,531</td>
<td>74,330</td>
</tr>
<tr>
<td>Subtotal</td>
<td>874,736</td>
<td>4,334,809</td>
<td>8,817,349</td>
</tr>
<tr>
<td>Total (hp-hr/yr)</td>
<td>1,642,878</td>
<td>8,138,303</td>
<td>17,166,675</td>
</tr>
<tr>
<td>Total (10E12) BTU/yr</td>
<td>0.0042</td>
<td>0.0207</td>
<td>.0437</td>
</tr>
</tbody>
</table>
Figure 9. Distribution of Yearly Energy Consumption
Figure 10. Scale Effects of Energy Consumption
the distribution of energy consumption is roughly half farm and half conversion facility.

The last line in Table 25 details the total process energy consumption in BTUs. This data reveals that the proposed process should consume about 4.3 percent of the yearly gas production. Chapter III's assumptions were quite broad and may have introduced some errors but it is unlikely that those assumptions were off by 50 percent. Hence, even the worst case scenario would produce 10 times more energy than was consumed by the process.

Economics

Cost effectiveness is the major driving force for the implementation of any new idea. To be cost effective a project must provide a cash flow which will recover investment with interest during the equipment's useful life.

Capital investment by component is shown tabularly in Table 26 and graphically in Figure 11. The total investment required is 12, 24 and 37 million dollars for the 0.1, 0.5 and 1.0(10E12) BTU per year systems, respectively. Of these amounts about 2/3 is conversion related equipment while 1/3 is farm related equipment. Figure 11 was prepared by distributing indirect costs proportionately among the individual direct component costs. Examination of this figure shows that the percentage cost of site and gas preparation decreases with scale while the percentage of the leaching-bed digester and packed-bed digester increases with scale. Figure 12 plots the normalized cost of
### TABLE 26

**CAPITAL INVESTMENT**

<table>
<thead>
<tr>
<th>System Size</th>
<th>Description</th>
<th>(10E12)</th>
<th>(10E12)</th>
<th>(10E12)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>BTU/yr</td>
<td>BTU/yr</td>
<td>BTU/yr</td>
</tr>
<tr>
<td>0.1</td>
<td></td>
<td>0.5</td>
<td>1.0</td>
<td></td>
</tr>
<tr>
<td></td>
<td>$1,000</td>
<td>$1,000</td>
<td>$1,000</td>
<td></td>
</tr>
</tbody>
</table>

**Conversion Facility**

<table>
<thead>
<tr>
<th>Description</th>
<th>(10E12)</th>
<th>(10E12)</th>
<th>(10E12)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Site improvements</td>
<td>575</td>
<td>748</td>
<td>941</td>
</tr>
<tr>
<td>Leaching-bed digester</td>
<td>921</td>
<td>2,419</td>
<td>4,838</td>
</tr>
<tr>
<td>Packed-bed digester</td>
<td>191</td>
<td>729</td>
<td>1,296</td>
</tr>
<tr>
<td>Gas preparation equipment</td>
<td>1,492</td>
<td>1,941</td>
<td>2,670</td>
</tr>
<tr>
<td>Miscellaneous process</td>
<td>1,135</td>
<td>2,427</td>
<td>3,505</td>
</tr>
<tr>
<td>Instrument/electrical</td>
<td>748</td>
<td>1,503</td>
<td>2,462</td>
</tr>
<tr>
<td>Indirect costs</td>
<td>3,083</td>
<td>5,948</td>
<td>9,569</td>
</tr>
<tr>
<td><strong>Subtotal</strong></td>
<td>8,145</td>
<td>15,715</td>
<td>25,281</td>
</tr>
</tbody>
</table>

**Farm**

<table>
<thead>
<tr>
<th>Description</th>
<th>(10E12)</th>
<th>(10E12)</th>
<th>(10E12)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Site improvements</td>
<td>210</td>
<td>248</td>
<td>285</td>
</tr>
<tr>
<td>Winches and collection boom</td>
<td>910</td>
<td>2,128</td>
<td>3,245</td>
</tr>
<tr>
<td>Conveyor</td>
<td>681</td>
<td>1,483</td>
<td>1,958</td>
</tr>
<tr>
<td>Pump</td>
<td>81</td>
<td>85</td>
<td>87</td>
</tr>
<tr>
<td>Pipeline</td>
<td>104</td>
<td>410</td>
<td>524</td>
</tr>
<tr>
<td>Instrument/electrical</td>
<td>355</td>
<td>821</td>
<td>1,163</td>
</tr>
<tr>
<td>Indirect</td>
<td>1,426</td>
<td>3,152</td>
<td>4,423</td>
</tr>
<tr>
<td><strong>Subtotal</strong></td>
<td>3,767</td>
<td>8,327</td>
<td>11,685</td>
</tr>
</tbody>
</table>

**Total**                             | 11,912  | 24,042  | 36,966  |
<table>
<thead>
<tr>
<th>Conversion Site 4.1%</th>
<th>LEACH-BED DIGESTER 21.1%</th>
<th>PACKED-BED DIGIT 5.6%</th>
<th>GAS PREP EQUIPMENT 11.6%</th>
<th>MISC PROCESS EQUIPMENT 15.3%</th>
<th>INSTRUMENT/ELECTRICAL 10.7%</th>
<th>FARM COMPONENTS</th>
</tr>
</thead>
<tbody>
<tr>
<td>7.8%</td>
<td>5.0%</td>
<td>12.4%</td>
<td>16.2%</td>
<td>2.6%</td>
<td>4.9%</td>
<td>0.1(10E12) BTU/yr</td>
</tr>
<tr>
<td>12.4%</td>
<td>16.2%</td>
<td>20.2%</td>
<td>13.0%</td>
<td>15.3%</td>
<td>16.2%</td>
<td>0.5(10E12) BTU/yr</td>
</tr>
<tr>
<td>2.6%</td>
<td>4.9%</td>
<td>10.1%</td>
<td>10.1%</td>
<td>10.1%</td>
<td>1.7%</td>
<td>1.0(10E12) BTU/yr</td>
</tr>
<tr>
<td>F-SITE 2.8%</td>
<td>WINCH/ COLLECTION BOOM 12.3%</td>
<td></td>
<td>14.2%</td>
<td>9.2%</td>
<td>9.9%</td>
<td></td>
</tr>
<tr>
<td>SHORE CONVEYOR 9.2%</td>
<td>PUMP 1.1%</td>
<td>PIPE 1.4%</td>
<td>IN/EL 4.8%</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Figure 11. Distribution of Capital Investment
Figure 12. Scale Effects of Capital Costs
the leaching-bed/packed-bed combination, the site improvements gas preparation combination and the total system costs. Here normalized cost is the cost of size X divided by cost of size 1.0 and the slope of these curves is an indicator of how fast cost increases with size. For a ten-fold increase in production, total system costs rise by a factor of 3.1, site improvements and gas preparation increase by a factor of 1.7 while digester costs increase by a factor of 5.5. Figure 13 shows the unit cost of capital investment. Since the curves start to level off near the 1.0(10E12) BTU per year system the range of plant sizes selected for analysis seem reasonable.

Operating and maintenance expenses by component are tabulated in Table 27 and illustrated graphically in Figure 14. The yearly cash flows are 1.0, 1.7 and 2.6 million dollars for the 0.1, 0.5 and 1.0 (10E12) BTU per systems, respectively. Generally 2/3 of these amounts is associated with the conversion facility while 1/3 is associated with the farm. Figure 14 details the exact distribution of expenses for each system size. Examination of this figure shows that the percentage of maintenance material expense is relatively constant, the percentage of personnel expenses decreases with increasing size while all the remaining components increase their percentage of total expense. All three systems are so small that manpower requirements are not affected by scale. The other components however, all increase with scale to some degree. Lime, fertilizer and electricity increase on approximately a one-to-one ratio. Materials increase about 3.1 times for a ten-fold increase in system capacity which is similar to
Figure 13. Unit Cost of Capital Investment
### TABLE 27

OPERATING AND MAINTENANCE EXPENSES

<table>
<thead>
<tr>
<th>Description</th>
<th>System Size</th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.1 (10E12)</td>
<td>0.5 (10E12)</td>
<td>1.0 (10E12)</td>
<td></td>
</tr>
<tr>
<td></td>
<td>BTU/yr</td>
<td>BTU/yr</td>
<td>BTU/yr</td>
<td></td>
</tr>
<tr>
<td>$1,000</td>
<td>($1,000)</td>
<td>($1,000)</td>
<td>($1,000)</td>
<td></td>
</tr>
</tbody>
</table>

#### Conversion Facility

- **Lime (@ $71/ton)**
  - 0.1: 34
  - 0.5: 171
  - 1.0: 342

- **Electricity (@ $13.50/10E6 BTU)**
  - 0.1: 26
  - 0.5: 131
  - 1.0: 287

- **Personnel**
  - 0.1: 398
  - 0.5: 381
  - 1.0: 398

- **Materials (@ 2.7% of capital)**
  - 0.1: 220
  - 0.5: 424
  - 1.0: 683

**Subtotal**
- 0.1: 678
- 0.5: 1,107
- 1.0: 1,710

#### Fann

- **Fertilizer (@ $.50/dry ton)**
  - 0.1: 5
  - 0.5: 24
  - 1.0: 48

- **Electricity (@ $13.50/10E6 BTU)**
  - 0.1: 30
  - 0.5: 149
  - 1.0: 303

- **Personnel**
  - 0.1: 184
  - 0.5: 202
  - 1.0: 184

- **Materials (@ 2.7% of capital)**
  - 0.1: 102
  - 0.5: 225
  - 1.0: 315

**Subtotal**
- 0.1: 321
- 0.5: 600
- 1.0: 850

**Total**
- 0.1: 999
- 0.5: 1,707
- 1.0: 2,560

* Total personnel expenses for 20 people @ $14.00/hr for 2,080 hours was distributed between fann and conversion based on percent capital investment.
Figure 14. Distribution of Operating and Maintenance Costs
the 2.6 increase for total expenses. When all of these trends are combined the result is that percentage personnel expense decreases with scale. Figure 15 shows these trends graphically by plotting the normalized cost of the expense elements and the total expense. Figure 16 indicates the unit cost of operating and maintenance expenses. Since the curves start to level off near the 1.0(10E12) BTU per year system, the range of plant sizes selected for analysis seems reasonable.

Equation 48 details the life cycle cost methodology used in this study. It determines the number of years to pay off the capital investment with interest. The solution procedure is trial and error until the present worth of future gas sales approximates the present worth sum of capital investment, operating maintenance and fuel (electricity). Table 28 shows several iterations for each size system and was developed using the assumptions outlined in Chapter III. It shows that the payoff period is in excess of 60 years for the 0.1(10E12) BTU per year system, 14 years for the 0.5(10E12) BTU per year system and 9 years for the 1.0(10E12) BTU per year system. These are the base case results and their significance should be judged by their sensitivity to Chapter III assumptions. Equation 48 is a very simplistic economic model which was developed to test the economic sensitivity of design assumptions. It is not suitable for determining the exact payoff period but rather to determine the effect on payoff period as a function of capital cost. The capital cost could be in error by as much as 25 percent. With the above perspective the following
Figure 15. Scale Effects of Operating and Maintenance Costs
Figure 16. Unit Cost of Operation and Maintenance
### TABLE 28

**LIFE CYCLE COST SUMMARY**  
(per Equation 48)

<table>
<thead>
<tr>
<th>Life (yr)</th>
<th>Gas Value ($10E6)</th>
<th>Total ($)</th>
<th>Elec. Cost ($10E6)</th>
<th>Const. Cost ($10E6)</th>
<th>Recurring Costs ($10E6)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Size X = 0.1(10E12) BTU/yr</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>20</td>
<td>10.8</td>
<td>22.7</td>
<td>.8</td>
<td>11.9</td>
<td>10.0</td>
</tr>
<tr>
<td>40</td>
<td>17.1</td>
<td>25.4</td>
<td>1.0</td>
<td>11.9</td>
<td>12.5</td>
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<td>60</td>
<td>20.7</td>
<td>26.2</td>
<td>1.1</td>
<td>11.9</td>
<td>13.2</td>
</tr>
<tr>
<td><strong>Size X = 0.5(10E12) BTU/yr</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>5</td>
<td>15.7</td>
<td>31.2</td>
<td>1.3</td>
<td>24.0</td>
<td>5.9</td>
</tr>
<tr>
<td>10</td>
<td>30.3</td>
<td>36.4</td>
<td>2.4</td>
<td>24.0</td>
<td>10.0</td>
</tr>
<tr>
<td>13</td>
<td>38.1</td>
<td>38.8</td>
<td>2.9</td>
<td>24.0</td>
<td>11.9</td>
</tr>
<tr>
<td>14</td>
<td>40.6 &gt;</td>
<td>39.6</td>
<td>3.1</td>
<td>24.0</td>
<td>12.5</td>
</tr>
<tr>
<td>15</td>
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<td>3.2</td>
<td>24.0</td>
<td>13.0</td>
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<tr>
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<td>15.1</td>
</tr>
<tr>
<td>40</td>
<td>85.5</td>
<td>48.0</td>
<td>5.0</td>
<td>24.0</td>
<td>19.0</td>
</tr>
<tr>
<td><strong>Size X = 1.0(10E12) BTU/yr</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
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<td>55.0</td>
<td>54.5</td>
<td>4.7</td>
<td>37.0</td>
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<tr>
<td>10</td>
<td>60.5</td>
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<td>170.9</td>
<td>73.8</td>
<td>10.6</td>
<td>37.0</td>
<td>26.2</td>
</tr>
</tbody>
</table>

**NOTES:**  
Gas Value = \( X \times \text{FSPWS}_{\text{gas}} / 10E6 \); per Table 10
Total = Elec. Cost + Const. Cost + Recurring Costs
Elec. Cost = \( E \times \text{FSPWS}_{\text{elec}} / 10E6 \); per Tables 9 and 25
Const. Cost = capital investment per Table 26
Recurring Costs = operating/maintenance expense minus electricity cost per Table 27
trends can be observed from Table 28. At the payoff period, the cash flow expenses (electricity and recurring costs) become a smaller percentage of total present worth as system size increases. Therefore, the analysis is more sensitive to capital cost assumptions at large sizes and conversely more sensitive to cash flow expenses at small sizes. Figure 17 shows these trends graphically. Normal service life for a process plant like the proposed system is 20 to 40 years. If a 40 year life is assumed then a 50 percent reduction in either cash flow expenses or capital costs would be required to make the 0.1(10E12) BTU per year system economically viable. This magnitude of error is detectable by this analysis and therefore it is unlikely that a 0.1(10E12) BTU per year system would ever be economical. Similarly, the 0.5(10E12) BTU per year system would require either a 156 percent increase in capital investment or cash flow expenses to not pay off in 40 years. This magnitude of error is also within this study's accuracy range and therefore it is unlikely that any system larger than 0.5(10E12) BTU per year would not be economical. The above discussion covers all assumptions associated with the production system when selling price of gas is $6.00 per million BTU. If the price of gas is $3.00 per million BTU (about the historical value of $3.13 [29]) then the economic analysis becomes much more sensitive. The 0.1(10E12) BTU per year plant is even more economically unattractive. The 0.5(10E12) BTU per year system will pay off in slightly over 40 years and a system approaching 1.0(10E12) BTU per year is required for economic viability. Now a 30 percent increase in
Figure 17. Distribution of Conversion Process' Present Worth at Payoff Period
capital investment or cash flow expenses would make the system uneconomical. This level of error approximates this study's accuracy so the 1.0(10E12) BTU per year system is probably still economical. An important aspect of larger payoff periods (resulting from lower gas sales revenues) is that cash flow expenses become increasingly more important to economic viability. For a 40 year life, total present worth is composed of equal portions of capital investment and cash flow dollars.

Recommendations

Although the results of this study indicate that methane from biomass is both economical and energy efficient, these findings are contingent on numerous assumptions and extrapolations of data. Many fundamental assumptions were necessary because the proposed process is unproven at the commercial scale and only limited data exists at the laboratory scale. To introduce this new technology it will be necessary to validate the assumptions through basic research. To be cost effective the basic research should be focused on those areas with the largest impact on cost and energy efficiency. This approach minimizes both research expenditures and the technology's introduction time. The priority research areas are as follows:

1. The leaching-bed digesters are the single most expensive components in the total system and their percentage cost increases as plant size increases. Therefore, the assumed
compaction factor and the 60-day operating schedule should be verified through pilot scale testing.

2. Another large cost item is the fann winch and boom collection system. The test data utilized to estimate drag forces was collected under conditions which differ significantly from the design conditions. The test area was 314 feet² while harvest area will exceed 160,000 feet². Towing tests should be conducted on large mats of water hyacinths to verify that conventional flat plate flow theory adequately predicts towing forces.

3. The compressor and the biomass supply pump account for over 90 percent of energy consumption. The critical assumptions associated with these two components are the final delivery pressure of the produced gas and the total pump head. A delivery pressure of 225 psig was utilized in this study but certain areas also have 20 and 40 psig distribution pressures [2]. Identifying those areas with both low distribution pressures and sufficient gas consumption (i.e. 0.5(10E12) BTU per year) would increase the technology's attractiveness. Total pump head was modeled after pulp slurries because of limited data. Frictional factors should be identified specifically for water hyacinth slurries and
settling tests should be conducted to verify that these slurries can be pumped long distances.

4. Operating and maintenance costs become more important to economic viability as payoff period increases. Since the payoff period is most sensitive to the future gas prices (which can not be precisely estimated), operating and maintenance costs are a potentially significant variable. Additional study to precisely define the required staffing levels and maintenance materials for the 0.5 and 1.0(10E12) BTU per year systems would help mitigate the consequences of lower gas prices as well as generally promote introduction of the technology.
V. SUMMARY AND CONCLUSIONS

This report presents the feasibility analysis of an unproven concept to produce pipeline quality methane gas from biomass. The biomass feedstock considered was water hyacinth. The conversion process studied was a two-stage biological process which utilizes leaching-bed digesters for the production of volatile acids and packed-bed digesters for the production of methane. The results indicate that the above concept should be both energy efficient (produce more energy than it consumes) and economically feasible (provide sufficient return on investment). This result is contingent on the verification and expansion of the rather small data base which forms the foundation of this analysis.

The proposed process is quite energy efficient since only 5 percent of the produced energy is consumed by the process. For a 0.1, 0.5 and 1.0(10E12) BTU per year facility the expected energy consumption of the process is 4.2, 20.7 and 43.7(10E9) BTU per year, respectively. Energy consumption is nearly a constant fraction of gas production because over 90 percent of consumption is related to biomass transfer and gas compression. Both of these are linear functions of gas production.

Capital costs are expected to be 12, 24 and 35(10E6) dollars (in $ 1981) for the 0.1, 0.5 and 1.0(10E12) BTU per year systems,
respectively. The leaching-bed digesters and the boom-winch collection systems are the first and second single largest cost components. Both of these components have considerable design uncertainty because data for their complete analysis is not available. Further research in these two areas has the potential to reduce capital cost significantly. This is particularly important because these two components do not experience substantial reductions in unit cost as plant size increases.

Operating/maintenance costs are expected to be 1.0, 1.7 and 2.6 (10E6) dollars (in $ 1981) for the 0.1, 0.5 and 1.0(10E12) BTU per year systems, respectively. Maintenance materials and personnel are the two major contributors to these costs and their precise evaluation will require a better understanding of the process than now exists.

When the capital investment and cash flow expenses are applied to a simple life cycle cost model, the payback periods are less than the proposed equipment life for system capacities greater than 0.5(10E12) BTU per year. Payback periods decrease with increasing system size and increase as the value of produced gas decreases. The most sensitive economic variable is the value of the produced gas, however even under the most pessimistic assumptions the 1.0(10E12) BTU per year facility would pay back in its lifetime.

In summary, the leaching-bed/packed-bed two-stage biological process to produce methane from biomass has the potential to provide pipeline quality substitute natural gas at a reasonable cost. To
资本化利用这一潜力，需要进行额外的研究来完善浸出床概念，提升绞车集水系统、设施人员需求和维护材料消耗。这些调查将会减少现有的设计不确定性，并进而提升投资者对概念的信心。
APPENDIX

Figure 18 presents a flow diagram for a leaching-bed/packed bed methane production system. It shows the stream flows, vessels and pumps of significance and their functions are as follows:

Stream Flows

Biomass feedstock supply stream [1] - This pipe carries water hyacinth slurry (suspended solids feedstock) from the biomass supply pump (P-9) at the lake shore area to the acid phase digester (T-1) at the conversion facility.

Volatile acid supply stream [2] - This pipe carries volatile acids (liquid feedstock) from the acid phase digester discharge pump (P-1) to the acid surge tank (T-2).

Lime slurry supply stream [3] - This pipe carries lime slurry (slaked lime solution of Ca(OH)₂) used for pH control from the lime slurry tank discharge pump (P-8) to both the the acid surge tank (T-2) and the recycle water tank (T-4).

Volatile acid feed stream [4] - This pipe carries pH modified volatile acids (liquid feedstock) from the acid surge tank (T-2) through the acid storage tank discharge pump (P-2) to the methane phase digester (T-3).
Figure 18. Flow Diagram - Leaching-Bed/Packed-Bed Reactor Process
MPD liquid effluent stream [5] - This pipe carries unconsumed volatile acids (liquid feedstock which failed to be converted to methane as it passed through the methane phase digester) from the methane phase digester (T-3) to the recycle water tank (T-4).

MPD gas effluent stream [6] - This pipe carries biogas (methane and carbon dioxide mixture) at 1 atmosphere from the methane phase digester (T-3) to the gas compressor. At the compressor the biogas' pressure is raised to 17 atmospheres and it is then forwarded to the gas clean-up system. Here the biogas is separated into one stream which is mostly methane and another stream which is mostly carbon dioxide.

APD liquid effluent stream [7] - This pipe carries volatile acids (liquid feedstock) from the acid phase digester's (T-1) sump to the acid phase digester discharge pump (P-1). Part of this flow will be directed to the acid phase digester's (T-1) recycle spray header while the balance of the flow will be directed to the acid surge tank (T-2) via the volatile acid supply stream [2].

Liquid waste stream [8] - This pipe carries waste water (and any associated suspended/disolved solids) generated by the conversion process from the acid phase digester's (T-1) recycle mixing station to the liquid waste storage tank (T-5).

Plant solid waste effluent stream [9] - This conveyor transports residual solids (the part of the solid feedstock which can not be readily converted to liquid feedstock) out of the acid
phase digester (T-1) to trucks. The trucks will transport this material to a disposal site away from the conversion facility. The disposal site could be either a conventional land fill or a local farming operation. The plant's solid waste effluent has the potential to be utilized as fertilizer.

Plant liquid waste effluent stream [10] - This pipe carries waste water from the liquid storage tank (T-5) to a disposal site away from the conversion facility through the liquid storage tank discharge pump (P-5). The disposal site could be a conventional spray concentration pond to evaporate the liquid or a local farming operation. The plant's liquid waste effluent has the potential to be utilized as irrigation water.

Lime slurry make-up water stream [11] - This pipe carries fresh water from the supply water stream [12] to the lime slaker where it is mixed with lime oxide (unslaked lime or CaO) to form a lime slurry (slaked lime solution of Ca(OH)$_2$).

Supply water stream [12] - This pipe carries fresh water from the well through the supply water pump (P-6) to both the lime slurry make-up water stream [11] and the recycle water tank (T-4).

**Vessels**

Acid phase digester T-1 - This vessel contains solid biomass which is received from the farm via the biomass feedstock supply stream [1]. Here the solid biomass is converted to volatile
acids (liquid feedstock by microbial action) and discharged via the APD liquid effluent stream \[ \text{7} \]. A small portion (residual solids) of the solid biomass will not be readily converted to liquid feedstock and is discharged via the plant solid waste effluent stream \[ \text{9} \].

**Acid surge tank T-2** - This vessel receives volatile acids (liquid feedstock) from the acid phase digester (T-1) via the volatile acid supply stream \[ \text{2} \] and the acid phase digester discharge pump (P-1). Here the volatile acids are pH modified by the addition of lime slurry via the lime slurry supply stream \[ \text{3} \] and are discharged via the volatile acid feed stream \[ \text{4} \]. A small amount of solids may settle out of the solution in this vessel and the acid surge tank solids pump (P-4) is required to carry these solids (along with a considerable amount of liquid) back to the acid phase digester (T-1). At the acid phase digester (T-1) the solids will be filtered out of the liquid by the biomass and gravel bed combination while the valuable liquid feedstock portion will be returned to volatile acid supply stream \[ \text{2} \]. This is necessary to prevent solids from reaching the methane phase digester (T-3) where they could damage the very sensitive microbial population. The microbial population in the acid phase digester (T-1) is not as sensitive and returning the solids to this location should not pose any operational problems.
Methane phase digester T-3 - This vessel receives pH modified volatile acids (liquid feedstock) from the acid surge tank (T-3) via the volatile acid feed stream [4] and the acid storage tank discharge pump (P-2). Here most of the liquid feedstock is converted to biogas (methane and carbon dioxide mixture by microbial action) and discharged to the gas compressor via the MPD gas effluent stream [6]. The portion of the liquid feedstock which is unconverted is discharged to the recycle water tank (T-4) via the MPD liquid effluent stream [5].

Recycle water tank T-4 - This vessel receives uncovered volatile acids (liquid feedstock) from the methane phase digester (T-3) via the MPD liquid effluent stream [5]. Here the liquid feedstock is conditioned as necessary for return to the acid phase digester (T-1). The exact requirements of this step are undefined at this time (due to limited understanding of the process) but provisions are provided to add fresh water from the supply water stream [12] and to add lime slurry from the lime slurry supply stream [3]. After conditioning, the tank effluent will be directed to the acid phase digester's (T-1) recycle mixing station by the recycle storage tank discharge pump (P-3). The majority of this liquid will be returned to the volatile acid supply stream [2] via the acid phase digester's (T-1) recycle spray header. However, a portion of this liquid (liquid waste stream [8]) will be expelled from the process via the liquid
storage tank (T-5) and the liquid storage tank discharge pump (P-5). This is necessary to prevent mass accumulation in the system because 95 percent of the biomass feedstock supply stream \[1\] is water.

**Liquid storage tank T-5** - This vessel receives the process' liquid waste from the acid phase digester's (T-1) recycle mixing station through the liquid waste stream \[3\]. Here the liquid waste is held in temporary storage until it is forwarded to a disposal site by the liquid storage tank discharge pump (P-5) through the plant liquid waste effluent stream \[4\].

**Lime slurry tank T-6** - This vessel receives lime slurry (slaked lime solution of Ca(OH)\(_2\)) from the lime slaker through the lime slaker discharge pump (P-7). Here the lime slurry is held in temporary storage until it is forwarded to the acid surge tank (T-2) or the recycle water tank (T-3) by the lime slurry tank discharge pump (P-8) through the lime slurry supply stream \[5\]. The lime slurry is utilized by the process to modify the pH of liquid feedstocks.

**Pumps**

**Acid phase digester discharge pump P-1** - This pump conveys volatile acids (liquid feedstock) from the acid phase digester's (T-1) sump to both the acid phase digester's (T-1) recycle spray header and the acid surge tank (T-2) through the APD liquid effluent stream \[6\] and the volatile acid supply stream \[7\].
Acid storage tank discharge pump P-2 - This pump conveys pH modified volatile acids (liquid feedstock) from the acid surge tank (T-2) to the methane phase digester (T-3) through the volatile acid feed stream [4].

Recycle storage tank discharge pump P-3 - This pump conveys the conditioned but unconverted volatile acids (liquid feedstock) from the recycle water tank (T-4) to the acid phase digester's (T-1) recycle mixing station.

Acid surge tank solids pump P-4 - This pump conveys a suspended solids solution from the bottom of the acid surge tank (T-2) to the acid phase digester (T-1) through the acid phase digester's (T-1) recycle mixing station.

Liquid storage tank discharge pump P-5 - This pump conveys waste water from the liquid storage tank (T-5) to a disposal site away from the conversion facility through the plant liquid waste effluent stream [10].

Supply water pump P-6 - This pump conveys fresh water from the well to both the lime slaker and the recycle water tank (T-4) through the lime slurry make-up water stream [11] and the supply water stream [12].

Lime slaker discharge pump P-7 - This pump conveys lime slurry (slaked lime solution of Ca(OH)₂) from the lime slaker to the lime slurry tank (T-6).
Lime slurry tank discharge pump P-8 - This pump conveys lime slurry (slaked lime solution of Ca(OH)\textsubscript{2}) from the lime slurry tank (T-6) to both the acid surge tank (T-2) and the recycle water tank (T-4) through the lime slurry supply stream [3].

Biomass supply pump P-9 - This pump conveys water hyacinth slurry (suspended solids feedstock) from the farm at the lake shore area to the acid phase digester (T-1) at the conversion facility through the biomass feedstock supply stream [4].


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