BIOLOGICAL CONVERSION OF BIOMASS TO METHANE
CORN STOVER STUDIES

Project Report
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ABSTRACT

A series of experiments was conducted to determine the performance characteristics of the methane fermentation process using corn stover as a substrate. Stover was obtained from the University of Illinois farms. This stover was processed through four parallel fermenters each having a capacity of 775 liters. A continuous feed system was employed to determine the conversion efficiency. The effluents from the fermentation unit were evaluated to determine the dewatering characteristics and the quality of the liquid and solid residues.

The biodegradability of corn stover is low. Data obtained at a fermentation temperature of $59 \pm 1^\circ\text{C}$ show that only 36 percent of the volatile solids are biodegradable. The first order rate constant for this conversion was found to be 0.25 day^{-1} . A mild thermochemical pretreatment step consisting of caustic (NaOH) concentration of 0.30 molar (5 g/100 g dry stover) and a temperature of 115°C for one hour increased the biodegradable fraction to 71 percent of the volatile solids.

The reactor slurries were easily dewatered by both vacuum filtration and centrifugation. In fact, a 10 mesh screen can capture a large percentage of the residual solids when thermochemical pretreatment is not practiced. Thermochemical pretreatment generates a much smaller particle size distribution requiring more efficient solids recovery systems. Solids capture with both the vacuum filter and centrifuge is poor unless chemical conditioning is employed.

Corn stover does not appear to be attractive economically at the present energy prices. Because of poor conversion efficiency, thermochemical pretreatment is required. At a chemical cost of \$154/tonne (\$140/ton), the

NaOH treatment adds approximately \$5.2/tonne to the cost of processing the stover. At a methane yield of $0.25 \text{ m}^3/\text{kg}$ of solids fed, this adds a total cost of $\$2/100 \text{ m}^3$ ($\$0.57/\text{MCF}$) for this process alone. Addition of stover acquisition costs ($\$20/\text{dry tonne}$ of stover), total processing costs without gas cleanup ($\$21/\text{tonne}$) and residue disposal ($\$3/\text{tonne}$ of wet cake), the cost of fuel gas would be in the neighborhood of $\$9.76/\text{GJ}$ ($\$10.30/10^6 \text{ BTU}$). This cost excludes all profit, taxes, etc. associated with private financing. Depending upon financing methods, tax incentives, etc., it may be necessary to add up to an additional $\$2.00/\text{GJ}$ to the cost of this fuel gas.

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INTRODUCTION

Recovery of energy from residual organic materials and from crops grown specifically for their energy content is part of the Fuels from Biomass Program. One of the processes for energy recovery is the methane fermentation system. In order to evaluate the methane yields from various organic materials a small pilot scale experimental system has been established. Construction of this facility was initiated in June of 1976, and it became operational in October 1976.

After the initial shakedown period, tests were conducted on the methane yields from beef feedlot manure. This initial substrate was selected to complement the work being conducted on other related projects. After completion of the manure studies, the substrate evaluated was corn stover collected from the University farms after the grain had been harvested. The results of these investigations are contained in this report.

One objective of the proposed research is to provide data on the methane yield for various operating conditions. These data will provide the necessary base from which economic feasibility studies can be conducted. Since a major factor in the economic evaluation is the yield of methane per unit of material processed, these data are essential for the above studies. Additional objectives include the evaluation of the benefits of substrate pretreatment and the evaluation of potential systems for processing and disposing of the fermentor slurry.

DESCRIPTION OF EXPERIMENTAL SYSTEM

Continuous Fermentation System

This methane fermentation system is housed in the Department of Civil Engineering's Dynamic Testing Lab. This building is located on the south campus of the University of Illinois at Urbana-Champaign. A floor plan of this facility is shown in Figure 1. In addition to a large test bay in which the test reactors were housed, this facility contained adequate laboratory, office and related space for the research staff.

The fermentation reactors consisted of four completely mixed stainless steel reactors with a total volume of 0.91 m^3 each. The operating level in each reactor was 0.775 m^3 . Heating was provided by circulating hot water from a water heater through external jackets on each tank. A temperature controller (TC) opened a solenoid valve (SV) to allow the hot water to circulate through the jacket when the temperature dropped below the set point. Figure 2 shows a schematic of one set of reactors. A second set operated in parallel.

The reactors were initially equipped with variable speed turbine mixers. The mixing speed could be varied from 20 to 120 rpm. Recording watt meters measured the power consumption on each mixer. The mixing power input could be varied from 0.1 kw to 1.0 kw. This is equivalent to mixing power input ranging from 0.13 to 1.3 kw per m^3 of slurry. Because of repeated failure of the variable speed drives, they were removed from the system in April 1977. The mixer speed was fixed at 75 rpm. The dimensions of the reactor and the associated mixer are shown in Figure 3.

The feed slurry was pumped into the fermenters with a progressing cavity pump (P1). This pump was connected to a time switch (TS) to

Dynamic Test Laboratory Plan

Figure 1. Floor Plan of Building for Housing the Experimental System

Figure 2. Schematic of One Set of Fermentors

Figure 3. Details of Reactor and Mixer

allow for the approximation of a continuous feed system by pumping a percentage of each hour. Effluent from the reactors was pumped by a similar type pump (P2) that was actuated by a level controller (LC). The effluent was collected in a holding tank (T1) for additional processing. Gas from the reactors was passed through wet test gas meters (M) for determining gas flow rate. The gas was sampled for gas analysis with a Fisher Gas Partitioner.

Feed Preparation

The organic materials used in this process have to be prepared for feed into the system. The initial step was one of size reduction. This was accomplished by a "granulator" manufactured by H. C. Davis Sons, Inc. This unit was equipped with four screens, 3.2 mm, 6.3 mm, 19 mm, and 39 mm. The organic material could be milled to either of these four desired sizes. In order to effectively process this material it had to be relatively dry. This also assisted in storage of significant quantities of feed material. The dry material did not decompose.

After size reduction, a predetermined weight of the feed material was added to the slurry tanks (ST). The dry feed was slurried with the desired amount of water. Either recycled liquor or fresh water could be used in the slurry tanks. After completely wetting the organic material, it was then discharged into mixed holding tanks (HT). This provided for storage of the slurry prior to pumping into the reactors. These mixed tanks also provided for a relative uniform feeding of material into the reactors since they contained approximately two days supply of feed slurry. These tanks were covered primarily for odor control.

Fermenter Operation

The loading rate on the reactors was determined by the concentration of organic solids in the feed slurry and the volume of slurry pumped per unit time. The predetermined loading rates were set by controlling the feed slurry concentration and varying the pumping time.

The performance of the fermentation system was analyzed on the basis of gas production. This was augmented with analysis of the reactor contents for total and volatile total solids, pH, alkalinity, and volatile acids. These data provided the basis for calculating the conversion efficiencies and methane yields. Each reactor was on an independent temperature control system so that the fermentation temperature could be varied as desired.

There were only two feed slurry holding tanks so two reactors were fed from each tank. However, by varying the throughput, the loading rate and retention time of each reactor could be varied. This arrangement essentially allowed each reactor to operate independent of the others. Therefore, it was possible to evaluate four operating conditions at one time.

Residue Processing and Disposal

The effluent from the reactors was collected in an effluent storage tank (T1). These storage tanks were connected to a progressing cavity pump (P3) that fed the centrifuge. The resulting centrifuge cake was collected for either further processing or disposal by landfill. The centrate was discharged into a holding tank. When centrate recycle was employed this liquor was used to make up the feed slurry. Without recycle this material was allowed to flow to the drain.

Two centrifuges were used. One was a Fletcher Mark III. This was a solid bowl basket centrifuge. This unit had a variable speed drive so that it is possible to evaluate the effect of centrifugal force on solids capture and cake moisture. This unit was augmented with a horizontal solid bowl conveyor centrifuge (Sharples P-600) that could be operated continuously. Larger quantities of slurry could be processed with the continuous unit than with the basket unit.

Provisions were made for applying a thermochemical treatment process for either the residue or the feed material. A mixed reactor capable of withstanding an internal pressure of 2 atm was connected to a steam generator. By adding the fibers to this reactor along with an appropriate amount of caustic and water it was possible to heat treat this material prior to fermentation. This permitted evaluation of the effect of thermochemical treatment of the solids on the fermentation characteristics.

In addition to the dewatering by centrifugation, the effluent was available for additional tests. This slurry can be dewatered by vacuum filtration and filter leaf tests were available to evaluate the slurry filtration characteristics. Also, the quality of the effluent liquid and solids was evaluated with respect to its nutrient value both as a possible animal feed material or as a soil additive. These evaluations were made with laboratory analyses rather than having actual field tests.

Fermenter Feed Stock

After completion of the fermentation studies using manure as the substrate, a series of tests were designed to evaluate the gas yield from corn stover. Corn stover is the residue remaining in the fields after the grain has been harvested. Because the grain is shelled from the cob, stover

contains the cob, stalk and leaves of the mature corn plant. Stover was collected from the corn fields adjacent to the laboratory. The material was collected mechanically and consisted of the above ground plant material that had either been cut or broken from the stem during grain harvest. Several wagon loads of stover were collected and stored in a covered storage area. Since the stover was not chopped during harvest, the particle size of the collected stover was large, with corn stalks up to approximately 20 cm in length. Prior to feeding to the fermenters, the corn stover was milled so that it would pass through a 6.3 mm (0.25 in.) screen initially and at a later date through a 3.2 mm (0.125 in.) screen.

The particle size distribution of the fermented solids plays a major role in determining the solids recovery potential. Also, the composition of these different size fractions will be important in evaluation of by-product recovery systems. The particle size determination was made using a wet sieving technique described in the Appendix B. A known volume of effluent slurry was passed through a series of sieves. The solids retained on each sieve were washed with a fixed volume of distilled water with the wash water being passed through each successive sieve. The sieves and the retained solids were oven dried and the dry weight of the retained solids determined. The solids passing through all of the sieves were determined by sampling the water collected from the sieves. Standard dry sieving techniques were utilized to determine the particle size distribution for the feed material milled through either the 3.2 or 6.3 mm screen.

Figure 4 shows the particle size distribution of corn stover milled on both the 3.2 mm and the 6.3 mm screen. The manure milled on the 6.3 mm screen is included as a comparison. Less than 5 percent of these samples passed through the 200 mesh sieve. Milling of the corn stover

Figure 4. Corn-Particle Size Distribution at Various Process Stages

actually resulted in particle size reductions as evidenced by the significant difference in the particle size distribution for the two different screen sizes. However, the manure clumps were simply broken apart. The screen size in the mill did not influence the particle size.

Particle size distribution curves for the feed slurry and digested slurry are also shown in Figure 4. The raw corn slurry solids were passed through the 3.2 mm screen. The total solids concentration of the slurry was 6.73 percent. The reactor was operated at a fermentation temperature of 58°C and a retention time of 15 days. The total solids content of the reactor effluent was 3.44 percent. These data show that only about 10 percent of the feed slurry and 15 percent of the reactor effluent solids were between 70 μ m and 1.98 mm in size. For the feed slurry, 42 percent of the solids were greater than 1.98 mm. For the dry stover, approximately one percent of the solids were greater than 1.98 mm. Clearly, the moisture absorbed by these solids caused significant expansion of the particles. For the fermented slurry, 63 percent of the solids were retained on the 10 mesh screen.

A significant portion of the total solids in the feed slurry (44 percent) passed through the 200 mesh sieve. This fraction includes both soluble solids and small suspended and colloidal solids. The reactor effluent contained a smaller percentage of solids passing through the 200 mesh screen. This will be discussed in more detail later.

Prior studies with feed lot manure did not require the addition of chemicals for pH control nor for bacterial nutrition. Corn stover is deficient in both nitrogen and phosphorus. Natural alkalinity does not form in sufficient quantity to maintain the pH above 6.6. Consequently, it was necessary to add nitrogen and phosphorus nutrients as well as lime

for pH control. The quantity of chemicals added depended upon the solids concentrations and the rate at which water was recycled within the process. Sufficient lime was added to the feed slurry to maintain the pH in the 6.7 to 6.8 range. Nitrogen addition was based on the need to maintain an ammonia nitrogen content of 200 mg/l in the fermentation tanks. Phosphorus was added at a rate equal to 0.2 of the nitrogen addition.

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RESULTS AND DISCUSSION

Corn Stover Fermentation

After completion of the manure fermentation studies at the end of November 1977, the substrate was changed to corn stover. During December of 1977, different techniques of processing and pumping of the stover were investigated. Significant problems were encountered in trying to pump the slurry of stover. Pumping of this slurry with total solids concentration of 5 to 6 percent was not possible. Examination of the Moyno pump showed extensive wear on the stators and rotors. The grit in the manure and small pieces of metal incorporated in the manure either from the feed lot or while in storage, did significant damage to these pumps. It was necessary to replace the rotors and stators.

The minimum size screen purchased with the mill was 6.3 mm (0.25 in.). After a few attempts at pumping this slurry milled through the above screen, it was clear that a smaller size particle would be necessary for ease of processing with the existing equipment. In addition to pumping problems, it was not possible to dewater the fermented slurry in the conveyor centrifuge. The small openings in the center feed mechanism and the cake discharge ports plugged frequently with the larger particles. A decision was made to mill through a 3.2 mm (0.125 in.) screen.

During the period of December 1, 1977 to January 14, 1978, the major effort was directed toward developing techniques for handling of the slurry and in modification of the pumping system so that the slurry could be handled. The fermenters received varying quantities of slurry and the microbial culture allowed to acclimate to the corn stover substrate.

A. Gas Production Rates

One of the primary objectives of these tests was to determine the efficiency of the methane fermentation system in converting the corn stover into methane. The four fermenters were operated under various conditions in order to obtain gas production data. Reactor 3 was operated as a control, with the retention time maintained in the 13 to 15 day range throughout the study. The average weekly loading rates and gas production data for this reactor are given in Table 1. The gas production is measured at room temperature and atmospheric pressure. Correction to standard conditions has not been applied. The corn stover feed to the system after 78035 was milled through the 3.2 mm screen. These data cover the period from January 21, 1978 (78021) through August 15, 1978 (78231).

These data show variations in gas production. The relatively low gas production after the week ending on 78084 was a result of the recycle of significant quantities of inert organic solids. Prior to Day 78021, the effluent from the fermenters was dewatered with the Sharples conveyor centrifuge, and the centrate recycled. Because of problems of plugging with the centrifuge, the reactor effluents were dewatered on a screen, and the liquid was then centrifuged prior to recycle. This practice was continued until Day 78083. At this time, the screened effluent was recycled directly to the mix tanks. A significant quantity of small organic solids was recycled, resulting in a higher volatile solids feed rate. These recycled solids were very stable, having been passed through the anaerobic fermentation process at least once. Consequently, the gas production per unit of volatile solids fed decreased. The effect of these recycle solids will be discussed in more detail.

Table 1. Control Reactor Performance

Week Ending	θ Days	Gas LPH	Feed Kg V.S./D	CH ₄	
				%	M ³ /Kg V.S. Fed
78021	14.5	11.4	1.26	55.3	0.12
28	15.2	12.5	1.50	53.4	0.11
35	14.4	17.1	1.87	53.8	0.12
42	14.6	17.8	1.79	54.9	0.13
49	13.4	20.2	2.00	50.9	0.12
56	14.6	20.1	2.08	51.0	0.12
63	14.3	21.6	2.21	52.6	0.12
70	13.9	24.1	2.16	53.1	0.14
77	15.0	24.5	2.40	53.9	0.13
84	15.0	27.6	2.33	53.0	0.15
92	13.2	15.1	2.24	53.4	0.09
99	13.8	16.1	1.81	52.9	0.11
106	14.2	13.1	1.83	52.8	0.10
113	14.0	12.0	2.01	52.8	0.07
120	13.2	15.7	2.00	54.8	0.10
127	14.0	13.9	2.09	56.5	0.09
134	14.9	15.8	1.86	49.4	0.10
141	13.6	15.1	2.12	54.6	0.09
148	12.0	15.1	2.08	57.4	0.10
155	13.4	13.4	1.95	61.4	0.10
162	14.6	12.9	2.10	59.8	0.09
169	14.9	11.9	2.07	59.0	0.08
176	13.9	15.3	1.93	58.7	0.11
183	12.8	15.1	1.98	55.1	0.10
190	14.7	11.9	1.63	59.4	0.10
203	13.3	13.4	2.02	58.7	0.09
210	13.8	11.6	1.69	55.9	0.09
217	14.1	10.3	1.45	57.5	0.10
224	13.8	11.7	1.63	57.4	0.10
231	14.8	10.7	1.47	55.1	0.10

The weather presented some major problems with operation and may be responsible for some of the data variation. A blizzard on Day 78026 (January 26, 1978) and again on Day 78043 (February 12, 1978) isolated the lab for 3 and 2 days, respectively. Of course, during this period, the reactors did not receive any substrate. From Day 78083 through Day 78086 (March 25-28, 1978), an ice storm disrupted the power to the lab. Fortunately, the ambient temperature did not drop much below freezing. Therefore, even though the heat in the building was off, no freezing problems were caused by the low temperatures.

The gas production from Reactor 1 is shown in Table 2. After an initial period of acclimation to corn stover, a thermochemical pretreatment step was introduced. Starting on Day 78014, milled corn stover was mixed with $\text{Ca}(\text{OH})_2$ (3 percent by weight of lime) and added to a pressure reactor. Water, generally recycled liquid, was added to this tank to increase the moisture to approximately 75 percent. Steam was then added to maintain a temperature of 110°C for approximately 4 hours. On Day 78058, the lime was increased to 6 percent by weight and on Day 78079 to 10 percent by weight. There is no apparent improvement in gas production as a result of this pretreatment. After heating for a 4-hour period, the pH of the 10 percent lime treatment was in the range of 9.1 to 9.3. Additional water was added to this tank to obtain the desired slurry concentration before transfer to the holding tank.

Reactor 2 received the same treatment as Reactor 1 (see Table 3). The only variable was a reduced retention time during part of the period. The feed particle size for these reactors was the same as that fed to the control. The thermochemical pretreatment was terminated on Day 78098. The feed to all reactors was the untreated corn stover and the retention

Table 2. Gas Production from Reactor #1

Week Ending	Days	Gas LPH	Feed Kg V.S./D	CH ₄	
				%	M ³ /Kg V.S. Fed
78021	14.7	12.8	1.24	56.4	0.139
28	13.9	13.7	1.24	53.8	0.143
35	13.7	11.1	1.10	54.6	0.132
42	13.7	12.8	1.28	54.5	0.131
49	13.1	13.8	1.62	53.4	0.109
56	13.7	11.5	1.62	53.2	0.091
63	14.4	14.0	1.56	53.2	0.115
70	14.3	17.4	2.12	51.7	0.102
77	14.6	19.5	2.23	55.3	0.116
84	14.0	20.7	2.64	55.7	0.105
92	14.5	26.9	2.88	52.9	0.118
99	13.9	30.0	2.54	54.2	0.154
106	7.2	17.4	3.93	53.3	0.057
111	7.2	18.4	4.02	52.7	0.058
120	7.1	17.1	3.82	55.4	0.071
127	7.4	23.7	3.99	52.5	0.075
134	8.3	23.7	4.00	51.8	0.074
141	12.9	48.4	2.67	54.8	0.238
148	14.1	41.4	2.25	58.2	0.257
155	14.2	35.2	2.04	61.0	0.253
162	14.2	24.4	2.02	58.1	0.168
169	13.5	22.8	2.05	56.2	0.150
176	14.4	22.8	1.69	56.3	0.182
183	13.9	15.9	1.52	57.5	0.144
190	14.2	11.6	1.41	55.6	0.110
203	14.2	26.8	1.56	55.8	0.230
210	13.9	26.0	1.34	60.0	0.279
217	14.0	23.2	1.36	58.6	0.240
224	14.8	26.3	1.47	57.1	0.245
231	14.5	22.7	1.22	57.7	0.258

Table 3. Gas Production from Reactor #2

Week Ending	Days	Gas LPH	Feed Kg V.S./D	CH ₄	
				%	M ³ /Kg V.S. Fed
78021	14.7	13.9	1.24	54.6	0.147
28	14.6	13.5	1.17	53.0	0.147
35	13.7	10.9	1.10	54.4	0.129
42	14.2	15.2	1.23	53.9	0.159
49	13.6	14.9	1.66	53.6	0.115
56	13.2	14.1	1.68	52.8	0.106
63	13.9	14.1	1.81	53.7	0.100
70	10.6	21.8	3.00	52.6	0.092
77	8.5	28.8	3.91	54.0	0.096
84	8.8	30.3	4.09	54.4	0.097
92	10.7	37.7	4.40	52.9	0.109
99	9.0	38.1	3.88	53.8	0.127
106	4.9	23.3	5.70	53.1	0.052
113	5.0	25.8	5.88	53.5	0.057
120	5.1	24.4	5.32	53.7	0.059
127	5.1	24.6	6.02	52.3	0.051
169	8.5	33.3	3.24	-	0.141
176	8.2	34.5	3.10	57.5	0.154
183	7.5	34.8	2.61	57.7	0.185
190	7.8	20.0	2.64	54.9	0.100
197	7.7	23.1	2.96	57.4	0.108
203	7.2	49.0	3.07	60.3	0.231
210	8.1	41.6	2.41	58.9	0.244
217	7.3	35.3	2.60	55.7	0.176
224	7.6	36.8	2.90	53.6	0.163
231	7.6	38.5	2.81	55.7	0.183

Table 4. Gas Production From Reactor #4

Week Ending	θ Days	Gas LPH	Feed Kg V.S./D	CH ₄	
				%	M ³ /Kg V.S. Fed
78021	14.6	12.5	1.26	53.6	0.132
28	14.8	13.0	1.53	52.6	0.107
35	14.3	16.9	1.88	53.5	0.115
42	13.9	16.3	1.90	54.5	0.112
49	13.3	22.4	2.08	51.1	0.132
56	13.4	20.3	2.30	51.4	0.108
63	13.0	19.6	2.51	51.9	0.097
70	13.7	17.2	2.27	50.1	0.091
77	13.9	24.8	2.56	53.7	0.125
106	3.7	21.2	7.25	50.9	0.036
113	3.9	25.4	7.45	54.3	0.044
120	3.9	30.2	6.76	54.6	0.059

Daily Performance Data at Short Retention Time

78114					
115	3.9	28.3	7.29	54.6	0.052
116	3.8	28.7	7.91	54.0	0.049
117	3.8	29.9	6.69	56.7	0.060
118	4.0	31.2	5.84	53.1	0.071
119	3.8	32.5	6.00	-	0.072
120	3.8	30.7	6.83	-	0.060

times were changed in preparation for obtaining some rate data.

The data for Reactor 4 are shown in Table 4. This reactor was initially intended to duplicate Reactor 3. However, excessive variations in the temperature caused erratic data. It was necessary to shut this reactor down and correct the temperature control problem. This reactor was reseeded and started again for the rate studies.

The methane production data for Reactors 1 through 3 during the period when lime was used in the thermochemical pretreatment of the feed slurry for Reactor 1 and 2 are summarized in Table 5. The effects of any

Table 5. Methane Production Rates

Control (3)		Reactor 1		Reactor 2	
θ (days)	m ³ /kg V.S.	θ (days)	m ³ /kg V.S.	θ (days)	m ³ /kg V.S.
Lime - 3%					
14.7	0.117	14.1	0.138	14.3	0.141
14.2*	0.123	13.5	0.110	13.7	0.127
Lime - 6%					
14.4	0.132	14.4	0.110	11.0	0.096
Lime - 10%					
14.0	0.117	14.1	0.126	9.5	0.111

* Size of milled stover reduced from 6.3 to 3.2 mm

pretreatments employed were insufficient to show any trends. There was some variation in the data. This variation was due to a number of factors, some of which have been previously discussed. However, any significant improvement in biodegradability would have been obvious. Approximately 30 percent of the volatile solids were being fermented. An increase in the biodegradability to 50 percent would have resulted in methane productions of approximately 0.2 m³ per kg volatile solids added. Clearly no such increase occurred.

In order to determine the effect of retention time on the gas production rates, a run was made using retention times ranging from 3.8 to 13.7 days. Although the reactors were designed to be fed continuously, it was not possible to do so with this corn stover. Continuous feed requires the feed pumps to operate on a time switch, pumping at short intervals during the day. Unfortunately, the stover slurry would not

pump without an initial flush with water. Once the slurry was pumping, there were no problems. Therefore, it was necessary to manually operate the feed pumps. The reactors were fed approximately 50, 100, 150, and 200 liters per day. The appropriate volume was fed in 50 liter quantities in order to reduce the surge effect at the high feed rate.

The effect of retention time on conversion efficiency is shown in Table 6. The four reactors were operated at various retention times as shown in this table. Reactor 3 had been operating for approximately 15 weeks at this retention time prior to collecting the data shown in Table 6 (see Table 1). These data were collected during the period from 78101 to 78129.

Reactor 1 was operated at approximately 7.8 day retention time from 78101 to 78136. The data in Table 6 were taken only from 78116 to 78136. This would represent a period when the reactor could be expected to have reached equilibrium. The retention time for Reactor 2 was set at 5 days on 78101 and continued until 78129. The data collected during the period from 78109 to 78129 were used in Table 6. Reactor 4 was operated

Table 6. Effect of Retention Time on Gas Production

Reactor No.	θ Days	Measured CH_4 $\text{m}^3/\text{kg V.S. Fed}$	Correction Factor	Adj. CH_4 $\text{m}^3/\text{kg V.S. Fed}$
4	3.8	0.046	1.48	0.068
2	5.0	0.056	1.48	0.083
1	7.8	0.073	1.38	0.101
3	13.7	0.090	1.43	0.129

at 200 liter per day feed rate, resulting in 3.8 day retention time. This feed rate was initiated on 78101 and the data shown in Table 6 were collected from 78109 to 78120.

During this period, the reactor effluents were screened to remove the coarse solids. Fine suspended solids passed through the screen and were incorporated in the recycle liquor used to slurry the corn stover being fed to the fermenters. The organic solids contained in this liquor had been fermented and were essentially biologically inert. These solids added to the quantity of volatile solids fed to the fermentation tanks, resulting in lower methane production rates as shown by the column labeled Measured CH_4 . This methane production was calculated from the total volatile solids fed to the fermenters.

In order to determine the impact of the recycle of these solids on gas production, sampling of the recycle liquor was initiated. The total volatile solids of the recycle liquor varied significantly, ranging from low values of 5 g/l to high values of 15 g/l. This variation was due to a number of factors, including how well the recycle storage tank was mixed prior to using the recycle liquor to slurry more corn stover. By measuring the recycle solids concentration and the volume of recycle added, the mass of the recycle solids added to the feed was calculated. This allowed for the calculation of the correction factor shown in Table 6. When this factor is applied to the measured methane production, an adjusted methane production rate is calculated. This is the gas production rate based on the volatile solids added as new corn stover.

This adjustment is valid as can be seen by comparing the gas production from the control reactor. During the period from 78050 through 78084, all of the recycle liquor was passed through the Sharples centrifuge.

The recycle liquor contained approximately 6.0 g/l of volatile total solids. The soluble solids, fixed and volatile, were about 4.5 g/l. The measured CH_4 production during this period averaged $0.132 \text{ m}^3/\text{kg}$ volatile solids fed with an average retention time of 14.6 days (see Table 1). The adjusted CH_4 production in Table 6 for Reactor 3 operating at a 13.7 day retention time is $0.129 \text{ m}^3/\text{kg}$ volatile solids fed. These data show that the mass of recycle volatile solids can significantly alter the measured gas production rate.

The adjusted gas data can be used to determine the kinetics of the fermentation. The first step is to determine the biodegradability of the substrate. This must be done in order to establish the S_0 , the initial substrate level. S_0 can be determined graphically by linearizing a plot of $S_0 - S$ ($\text{m}^3 \text{ CH}_4/\text{kg}$ V.S. fed) against the reciprocal of the retention time, θ . When θ approaches infinity, θ^{-1} approaches zero. Also, the substrate remaining, S , will also approach zero. Consequently, $S_0 - S$ will approach S_0 .

Figure 5 shows a plot of the log of $S_0 - S$, expressed as methane production, vs. the reciprocal of the retention time. The four points provide a reasonably good fit to a straight line, with a y intercept of $0.16 \text{ m}^3 \text{ CH}_4/\text{kg}$ volatile solids fed. If one assumes that corn stover is a carbohydrate composed primarily of hexoses, the methane production at 100 percent conversion of substrate would be $0.45 \text{ m}^3/\text{kg}$. Based on this maximum possible methane production and the observed S_0 value of $0.16 \text{ m}^3/\text{kg}$, the corn stover appears to be only 36 percent biodegradable for the test conditions employed ($59 \pm 1^\circ\text{C}$ fermentation temperature).

Using the above biodegradability factor and the adjusted gas production in Table 6, a rate constant can be determined on the basis that the gas production rate is first order with respect to biodegradable

Figure 5. Determination of Biodegradability of Corn Stover

substrate. The following equation applies:

$$S = \frac{S_0}{1 + K\theta} \quad (1)$$

This equation can be linearized in the following form:

$$\frac{S_0}{S} = K\theta + 1 \quad (2)$$

The slope of the line will be K. A least square computation for the data in Table 6, including a value of $S_0/S = 1$ when $\theta = 0$, yield a slope of 0.25 day^{-1} . Therefore, K is equal to 0.25 day^{-1} . This constant can be used to predict gas production rates and solids destruction at various retention times. This constant applies for 58-60°C fermentation temperatures. A plot of S_0/S vs. θ is shown in Figure 6.

B. Effect of Thermochemical Pretreatment

As reported previously, the use of lime in the thermochemical pretreatment process was ineffective in improving the methane production. The research at Stanford (Gossett *et al.*, 1976) has shown that sodium hydroxide can be an effective agent in the enhancement of the biodegradability of the organics. A series of tests was undertaken to evaluate this pretreatment technique in a larger scale system.

The milled dry corn stover was mixed with sodium hydroxide at the rate shown in Table 7. Tap water was added at the rate of 4 kg per kg of dry stover. With a caustic addition of 5 g/100 g of dry stover, the caustic concentration in this mixture was approximately 0.30 molar. After thorough mixing of this paste in the pressure reactor, the pH at the above

Figure 6. First Order Rate Constant Determination

Table 7. Effect of NaOH Pretreatment of Corn Stover

Week Ending	# Days	Gas LPH	Feed Kg V.S./m ³ -D	CH ₄		NaOH % Dry Solids
				%	M ³ /Kg V.S. Fed	
<u>REACTOR 1</u>						
134	8.3	23.7	5.2	51.8	0.07	0
141	12.9	48.4	3.4	54.8	0.24	5.0
148	14.1	41.4	2.9	58.2	0.26	5.0
155	14.2	35.2	2.6	61.0	0.25	5.0
162	14.2	24.4	2.6	58.1	0.17	3.0
169	13.5	22.8	2.6	56.2	0.15	3.0
176	14.4	22.8	2.3	56.3	0.18	3.0
183	13.9	15.9	2.0	57.5	0.14	3.0*
190	14.2	11.6	1.8	55.6	0.11	3.0*
203	14.2	26.8	2.0	55.8	0.23	5.0
210	13.9	26.0	1.7	60.0	0.28	5.0
217	14.0	23.2	1.8	58.6	0.24	5.0
224	14.8	26.3	1.9	57.1	0.25	5.0
231	14.5	22.7	1.6	57.7	0.26	5.0
<u>REACTOR 2</u>						
162	14.4	19.7	3.1	57.7	0.11	3.0
169	8.5	33.3	4.2	-	0.14	3.0
176	8.2	34.5	4.0	57.5	0.15	3.0
183	7.5	34.8	3.4	57.7	0.18	3.0*
190	7.8	20.0	3.4	54.9	0.10	3.0*
203	7.2	49.1	4.0	60.3	0.23	5.0
210	8.1	41.6	3.1	58.9	0.24	5.0
217	7.3	34.2	3.4	55.9	0.18	5.0
224	7.6	36.7	3.7	53.6	0.16	5.0
231	7.6	38.5	3.6	55.7	0.18	5.0

* NaOH + Ca(OCl)₂ added to pretreatment step

caustic addition was 12.1 to 12.2. This mixture was heated for 4 hours at 115°C. After the heat treatment, the pH decreased to approximately 9.5. This paste was then diluted with recycled water to about 6 percent solids and pumped to the mixing tank for feed to the reactors.

The results of this pretreatment are given in Table 7. A caustic dosage of 5 g/100 g dry solids resulted in a significant increase in methane production. The data for 78137 through 78157 for Reactor 1 show a methane production rate of 0.25 m³ per kg of volatile solids fed with an average retention time in the fermenters of 13.7 days. This pretreatment was repeated again in Reactor 1 during the period from 78197 through 78231. With an average retention time of 14.2 days, the methane production was 0.25 m³ per kg of volatile solids fed. If the rate constant for treated and untreated stover remains the same, the increase in biodegradability can be estimated from these data and the following equation:

$$S_0 - S = 0.25 \frac{\text{m}^3}{\text{kg}} = S_0 \left(\frac{K_D}{1 + K_D} \right) \quad (3)$$

The calculated value for S_0 is 0.32 m³ CH₄ per kg volatile solids fed. Based on an ultimate gas production of 0.45 m³ CH₄ per kg volatile solids destroyed, the biodegradability of the corn stover was increased from 36 percent to 71 percent. The gas production in the control reactor is given in Table 1.

The above calculation applied to the data collected from Reactor 2 during the period from 78197 through 78231 yield an S_0 value of 0.30 m³ CH₄ per kg volatile solids fed. This would result in a 67 percent biodegradability. There was substantial variation in the four weeks of gas production. Therefore, one should not place too much confidence in these latter numbers.

Lower caustic dosages were investigated. The pretreatment for Reactors 1 and 2 during the period 78158 through 78178 was 3 g NaOH per 100 g dry solids. The initial pH of the mixed stover paste was in the range of 11.8 to 12.0. After the 4 hour heating period, the pH was approximately 9.0. While there did not appear to be a major change in the initial and final pH at this level of NaOH, there was a substantial reduction in the gas production. The methane production in Reactor 1 operating at a 14.0 day retention time was $0.17 \text{ m}^3 \text{ CH}_4$ per kg volatile solids fed. Reactor 2, operating at a retention time of 8.4 days, had a methane production rate of 0.145 m^3 per kg volatile solids fed. These data clearly show that significant caustic dosages are required to enhance the biodegradability of the corn stover.

An attempt was made to see if an oxidant such as chlorine would have any effect on the conversion efficiency. Calcium hypochlorite and sodium hydroxide were added to the dry stover each at the rate of 3 g per 100 g. The two week period shown in Table 3 indicates a substantial reduction in gas production. No attempt was made to determine what caused this reduction. The experiment was not continued as it was obvious that under these test conditions, gas production decreased.

Slurry Particle Size Distribution and Composition

The particle size distribution of corn stover depended upon the degree of size reduction obtained during the milling step. This was not the case with the feedlot manure. Prior size reduction occurred in the animal feed preparation as well as in rumination by the animal. Figure 4 illustrated the typical size distribution for the feedlot manure and the corn stover milled through two different size screens. The data in Table 8 were extracted from Figure 4.

Table 8. Particle Size Distribution of Dry Feed Material

Material	Screen Size - mm	Particle Size Exceeded by Stated Percentage of Total Particle Weight		
		80	50	20
Stover	3.2	0.23 mm	0.50 mm	0.84 mm
Stover	6.3	0.48 mm	1.10 mm	1.80 mm
Manure	6.3	0.25 mm	0.61 mm	1.30 mm

Milling of the corn stover through the 3.2 mm screen resulted in a significantly smaller particle than obtained with the 6.3 mm screen. In all three size categories in Table 8, the 6.3 mm screen produced particles approximately 2.1 times as large as those obtained with the smaller screen. However, the manure milled through the 6.3 mm screen yielded a particle size comparable to that obtained when the stover is passed through the small screen.

The importance of particle size reduction in the processing of corn stover has not been clearly defined to date. Since this material does not receive any significant size reduction in harvesting, it is probable that some milling will be required to effect adequate conversion efficiencies. Based on the gas production data, as discussed earlier, there is no advantage in milling the stover to pass the 3.2 mm screen as compared to the 6.3 mm screen. The maximum particle size permitted for efficient gas production has not been determined from these studies.

Analysis of the corn stover was conducted to determine the nutrient level and the cellulose content. The analysis of six separate samples of corn stover yielded the results shown in Table 9. If only the carbon in the total cellulose is considered to be available, the carbon to nitrogen

Table 9. Composition of Corn Stover

Constituent	% Total Solids	% Volatile Solids
TKN	0.63	0.83
Phosphorus	0.11	0.15
Potassium	0.81	1.07
Total Cellulose	46.6	58.4
Lignin	23.5	29.4
Ash	20.2	-

ratio is 33 to 1. This should be adequate to support microbial activity. However, it was necessary to add nitrogen in order to maintain approximately 200 mg/l of ammonia nitrogen in the reactor. While significant nitrogen appeared to be present in the stover, it apparently was not available. The additional nitrogen requirements depended upon the amount of centrate recycle employed. Phosphorus was added, as K_2HPO_4 , to supplement the phosphorus available. This also resulted in an increase in the potassium level of the slurries. Consequently, the quantity of N, P and K in the slurries will be a function of the amount of chemicals added. However, the distribution of these plant nutrients in the slurry may be of significance.

When the stover that had been milled through the 3.2 mm screen was slurried with water, the particle size distribution changed significantly. The data in Table 10 show that the total solids were either large or small. In the feed slurry, 76.1 percent of the total solids were larger than 1.98 mm. For the remaining 23.9 percent, 21.1 percent were smaller than 74 μ m. This fraction includes the soluble solids as well as small colloidal particles.

Table 10. Slurry Particle Size Distribution - No Thermochemical Pretreatment

Sieve Size	Percentage of Slurry Solids Retained		
	Total Solids	Volatile Total Solids	Suspended*
Feed Slurry (4 replicates)			
10	76.1	76.5	96.3
20	76.8	-	97.4
50	77.8	-	98.4
100	78.4	-	99.4
200	78.9	79.4	100.0
Pan	100.0	100.0	-
Reactor Effluent, $t = 14.7$ days, Temp = 58°C (4 replicates)			
10	63.2	65.2	89.9
20	65.6	-	93.3
50	67.3	-	95.6
100	68.7	-	97.6
200	70.3	72.6	100.0
Pan	100.0	100.0	-

* Suspended solids are considered to be those solids retained on the 200 mesh sieve

Thermochemical pretreatment of the corn stover would be expected to shift the particle size distribution. As shown by the data in Table 11, there was a relatively minor shift in the solids distribution when the low NaOH dosage (3 g/100 g) was used. This pretreatment step resulted in a marginal increase in gas production. When a higher dosage of NaOH (5 g/100 g) was employed, there appeared to be a significant shift in the particle size distribution as shown in Table 12. The solids passing the 200 mesh sieve (74 μ m) increased from 21.1 percent with no thermochemical pretreatment to 34.1 percent at the higher NaOH level. The solids retained on the 10 mesh screen decreased because of the above, but also because the solids in the size range between the 10 and 200 mesh sieve increased.

Table 11. Slurry Particle Size Distribution - 3 Percent NaOH Pretreatment

Sieve Size	Percentage of Slurry Solids Retained		
	Total Solids	Volatile Total Solids	Suspended* Solids
Feed Slurry (1 replicate)			
10	73.7	66.3	96.3
20	74.7	-	97.8
50	75.8	-	99.1
100	72.6	-	99.5
200	100.0	100.0	-
Reactor Effluent, $\theta = 7.75$ days, Temp = 58°C (3 replicates)			
10	63.7	67.3	91.7
20	66.3	-	95.4
50	67.5	-	97.2
100	68.4	-	98.5
200	69.4	73.4	100.0
Pan	100.0	100.0	-
Reactor Effluent, $\theta = 14.7$ days, Temp = 58°C (2 replicates)			
10	36.3	34.4	75.8
20	42.8	-	89.5
50	44.7	-	93.6
100	46.3	-	97.1
200	47.6	45.1	100.0
Pan	100.0	100.0	-

* Suspended solids are considered to be those solids retained on the 200 mesh sieve

Based on these analyses, it would appear that the particle size was changed by the thermochemical pretreatment. However, this particle size change does not appear to be enough to account for the very substantial increase in gas production obtained with this pretreatment.

The action of the microorganisms during fermentation tended to reduce the particle size. As can be seen in Table 10, there was a significant

Table 12. Slurry Particle Size Distribution - 5 Percent NaOH Pretreatment

Sieve Size	Percentage of Slurry Solids Retained		
	Total Solids	Volatile Total Solids	Suspended Solids
Feed Slurry (2 replicates)			
10	60.0	59.5	91.1
20	62.1	-	94.4
50	64.6	-	98.0
100	65.3	-	99.1
200	65.9	65.4	100.0
Pan	100.0	100.0	-
Reactor Effluent, $\theta = 5.2$ days, Temp = 58°C (2 replicates)			
10	56.3	59.1	84.0
20	58.8	-	87.9
50	63.6	-	95.2
100	65.2	-	97.5
200	66.8	70.1	100.0
Pan	100.0	100.0	-
Reactor Effluent, $\theta = 14.7$ days, Temp = 58°C (1 replicate)			
10	27.8	27.8	71.8
20	33.4	-	88.6
50	36.0	-	95.2
100	37.0	-	97.9
200	37.8	37.8	100.0
Pan	100.0	100.0	-

* Suspended solids are considered to be those solids retained on the 200 mesh sieve

decrease in the solids retained by the 10 mesh sieve, decreasing from 76.1 percent in the feed slurry to 63.2 percent in the reactor slurry. This effect is also shown in the "suspended" solids. In the feed slurry, 96.3 percent of this fraction was retained by the 10 mesh sieve. In the reactor slurry, the percentage was only 89.9. These data show that the fermentation increases both the soluble solids and the weight of solids found in the smaller size range.

The soluble solids in the reactor slurries accounted for about one-half of the solids passing the 200 mesh sieve. When feeding untreated stover, a reactor effluent containing 36.0 g/l of total solids contained 4.8 g/l of soluble solids. The soluble solids accounted for approximately 14 percent of the total solids. As can be seen from the data in Table 10, approximately 30 percent of the total solids passed the 200 mesh sieve. This is a significant quantity of very small insoluble solids that will be difficult to remove from the filtrate/centrate.

The caustic pretreatment (3 percent NaOH) did not greatly alter the feed slurry particle size distribution. However, as shown by the data in Table 11, the microorganisms do change this size distribution, especially at longer retention times. The reactor with the 7.75 day retention time did show a significant reduction in the larger size particles. However, the "suspended" solids distribution was not substantially different from the feed slurry.

The effect of the longer retention time on the particle size distribution was very pronounced. Over one-half, 52.4 percent, of the total solids passed the 200 mesh sieve. Samples of the effluent from this reactor contained 35.1 g/l of total solids and 6.5 g/l of soluble solids. Therefore, less than 20 percent of the total solids were soluble. The 30+ percent of the total solids passing the 200 mesh sieve were fine suspended solids.

Solids recovery from this effluent will be much more expensive. In the reactor receiving untreated stover, screening with a 10 mesh screen would recover 63 percent of the total solids and 90 percent of the solids that are greater than 74 μm size (see Table 10). Similar results would be obtained with the pretreated feed at a 7.75 day retention time. However,

at the 14.7 day retention time, only 36.3 percent of the total solids and 75.8 percent of the "suspended" solids can be captured with the 10 mesh screen.

An increase in the NaOH dosage used in the pretreatment only increases the above problem. The feed slurry at the 5 percent NaOH dosage exhibits a smaller particle size, with 34.1 percent of the total solids passing the 200 mesh sieve. Processing at a 5.2 day retention time changes the total solids only a little. More of the solids are retained on the intermediate sieves. Therefore, the "suspended" solids capture obtained with a 10 mesh screen would be less than obtained in the effluents shown in the previous tables.

The 14.7 day retention time shows a very dramatic change in the particle size distribution. For the total solids, 62.2 percent pass the 200 mesh sieve, and only 27.8 percent of these solids can be removed with the 10 mesh sieve. In addition, there is a wider distribution in the "suspended" solids size. Consequently, solids recovery from these effluents will require an investment in equipment and probably chemicals of a much greater magnitude than the simple screens. The value of these solids must be carefully assayed before investing in the solids recovery system.

Analysis of the slurry solids to determine the composition of various fractions and changes associated with different treatment schemes were conducted. The solids collected on the sieves during the particle size distribution tests were used to determine the composition of these solids. The solids retained by sieves other than the 10 mesh were of insufficient quantity to use in the analyses. Therefore, only those solids retained on the 10 mesh sieve and those captured in the pan were of sufficient quantity to use for analysis. The results are shown in Table 13.

Table 13. Composition of Various Size Fractions

Particle Size	% Total Slurry Solids	Composition - % of Size Fraction					
		Holo-Cellulose	Nitrogen		K	P	Ash
		TKN	Organic				
Feed Slurry, No Pretreatment							
>1.98 mm	76.1	50.3	0.61	0.37	0.78	0.12	18.5
< 74 μ m	21.1	36.1	2.07	1.02	5.40	0.48	20.8
Reactor Effluent, No Pretreatment, θ = 14.7 Days							
>1.98 mm	63.2	53.0	0.65	0.39	0.26	0.10	23.6
< 74 μ m	29.7	31.3	1.76	0.83	5.04	0.48	31.5
Feed Slurry, 3% NaOH Pretreatment							
>1.98 mm	73.7	49.6	0.55	0.47	1.20	0.13	37.2
< 74 μ m	23.5	22.5	2.01	1.51	4.84	0.48	7.2
Reactor Effluent, 3% NaOH Pretreatment, θ = 7.75 Days							
>1.98 mm	63.7	49.2	0.61	0.37	0.26	0.09	19.4
< 74 μ m	30.6	24.8	2.08	1.14	4.53	0.41	33.7
Reactor Effluent, 3% NaOH Pretreatment, θ = 14.7 Days							
>1.98 mm	36.3	40.6	0.49	0.27	0.29	0.13	32.6
< 74 μ m	52.4	21.0	2.42	1.72	5.61	0.45	25.3
Feed Slurry, 5% NaOH Pretreatment							
>1.98 mm	60.0	52.5	0.37	0.28	0.54	0.10	20.0
< 74 μ m	34.1	29.7	1.25	0.83	5.07	0.42	18.3
Reactor Effluent, 5% NaOH Pretreatment, θ = 5.2 Days							
>1.98 mm	56.3	45.2	0.36	0.39	0.12	0.05	26.2
< 74 μ m	33.2	23.9	1.75	0.59	3.91	0.32	36.9
Effluent, 5% NaOH Pretreatment, θ = 14.7 Days							
>1.98 mm	27.8	23.5	1.05	0.94	0.18	0.15	52.8
< 74 μ m	62.2	32.0	2.46	1.06	4.01	0.19	52.9

From these data, one can see that the phosphorus content is very low in both size fractions. The phosphorus added to the system was soluble and, except for that utilized by the microorganisms, would be expected to be found in the soluble solids fraction. In general, the phosphorus content of larger size fractions of these slurries was about the same as the original corn stover (see Table 9). While a significant portion of the phosphorus is associated with the larger particles, it is probably unavailable. Certainly the soluble phosphorus in the liquid fraction would be more available for plant nutrients. A similar observation can be made for the potassium except that a much greater proportion of this ion was present in the liquid fraction.

The nitrogen data also show that most of the nitrogen is found in the fraction passing the 200 mesh sieve. Ammonia added for microbial nutrition is soluble and accounts for approximately one-half of the nitrogen found in this size fraction. The organic nitrogen content of those small size particles is significantly higher than that found in the larger size fraction. This would suggest that much of the bacterial mass is present in the particle size fraction passing the 200 mesh sieve. Any protein recovery scheme would necessarily concentrate on this size fraction. It is not obvious that this protein concentration is worth recovery. For example, the organic nitrogen content in the reactor effluent solids ($\theta = 14.7$ days, 3 percent NaOH pretreatment) in Table 13 was only 0.7 percent of the total solids in the small size fraction. For the 52.4 percent of the total solids contained in this size fraction, 64 percent are insoluble solids. Therefore, the maximum protein content ($6.25 \times$ organic nitrogen) of these insoluble solids would be 6.8 percent. A similar analysis for

any of the effluents would show that the potential for protein recovery from these slurries is marginal.

The change in the total cellulose composition of the solids in the different slurries reflects the conversion efficiency achieved. For those reactors receiving untreated stover, the cellulose content of the various size fractions did not change significantly. As the conversion efficiency increased, as was the case for the reactor receiving stover pretreated with 5 percent NaOH and operated at a retention time of 14.7 days, the cellulose content of the solids decreased substantially. This was especially true for the large size fraction.

Vacuum Filtration Studies

The objective of the vacuum filtration tests was twofold. The first objective was to develop a set of design parameters which will allow for sizing an efficient vacuum filtration unit to dewater the various slurries under investigation. The second objective was the development of several cause and effect relationships to aid in selecting and altering experimentation in order to achieve the most useful results. The experimental parameters can be divided into three categories: control variables, independent variables, and dependent variables. The control variables involve the condition or state of the sludge being investigated. Due to the dynamic test conditions of the total fermentation system, the incoming sludge is monitored to insure that the battery of tests is being conducted on a sludge of constant nature. The independent variables are those which can be systematically varied to ascertain the filtration characteristics of the slurries. The dependent variables are those which reflect the result of changing the independent variables. Table 14 outlines the three variable types.

Table 14. Deliniation of Control, Independent and Dependent Variables

Control	Independent	Dependent
Slurry Total Solids	Coagulant Type	Filtrate Total Solids
Slurry Particle Distribution	Coagulant Dose	Filtrate Dissolved Solids
	Polymer Type	Filtrate Suspended Solids
Slurry pH	Polymer Dose	Solids Capture
Slurry Temperature	Form Time	Filter Loading Rate
	Dry Time	
	Form Pressure	
	Drying Pressure	
	Filter Media Type	

The dependent variables are a measure of the process performance. This will determine the utility of this unit process for slurry dewatering. The cake solids, which defines the ratio of dry solids and water in the cake, are important should incineration of the cake be desired. If the cake contains 30+ percent solids by weight, combustion may proceed without the aid of auxiliary fuel. Or if the sludge is to be landfilled, the higher the cake solids, the lower the weight of wet cake required to be transported to the landfill site.

The solids capture is the ratio of the suspended solids retained on the filter to those fed to the filter. If recovery in the solid phase is the primary interest a high solids capture value is necessary. If the main purpose is to recycle the liquid fraction, a high solids capture value might be desired to avoid a build up of refractory solids in the fermenter.

The three filtrate related variables, total solids, settleable solids, and dissolved solids, are also needed for evaluation of liquid recycle possibilities. These parameters, along with chemical analysis of these solid fractions, determine the amount of nutrients and alkalinity available for recycle.

The filter loading rate is a design parameter derived from the information revealed in the filter leaf test. It intergrates the information into a loading parameter for the continuous vacuum filter system. Loadings are expressed as weight of solids processed per area of filter per time.

A laboratory filter leaf test procedure was used for the evaluation of the filtration characteristics of the reactor effluent slurries. This procedure employs a section of filter cloth that is commercially available for installation on a vacuum filter. Figure 7 shows a schematic of this apparatus. The test cloth is subjected to the same sequence of operations encountered in an operating vacuum filter. A detailed discussion of this experimental procedure is presented in Appendix B.

Evaluation of Filtration Characteristics

Vacuum filtration studies were conducted on digested corn stover during the period which started February 2, 1978, and ended July 26, 1978. Effluent slurries were taken from Reactor 3 and were used in all studies in which the stover was not pretreated. This unit served as a control, operating with a retention time of 14.7 days. The average total solids concentration of the effluent slurries for the thirty-seven days of testing was 4.59 percent.

Figure 7. Filter Leaf Test Apparatus

Various sludge conditioners were employed throughout these studies. Lump form ferric chloride with 6 waters of hydration was used in many instances. The ferric chloride was maintained in a stock solution at a concentration of 100 g/l. Two anionic polymers, Dow Purifloc A-22 and A-23 were also employed. Stock solutions were maintained at a concentration of 6 g/l. In addition, two cationic polymers C-31 and C-43 were used in stock solutions of 10 g/l. The concentrations of the stock solutions are expressed as the weight of chemical per final volume of solution. Also, the chemical dosages appearing in the following graphs are expressed as grams of conditioner per initial volume of slurry rather than final volume after chemical addition.

The alkalinity of the fermented corn stover slurry averaged about 1300 mg/l as CaCO_3 during the period of study. This value was much lower than the alkalinity encountered during the studies on cow manure. Consequently, the pH of the corn stover slurry was easier to depress with ferric salt addition. The relationship between pH depression of the corn slurry and ferric chloride addition is given in Table 15.

There are a number of other test parameters which are important to mention. Unless otherwise specified, the form pressure used in these tests was 16 inches of mercury. The initial drying pressure used was 15 inches of mercury. The drying pressure typically decreased during the drying cycle due to a loss of vacuum associated with the movement of air through the filter cake. In instances where the form pressure was not 16 inches of mercury, the initial drying pressure was 1 inch of mercury less than the form pressure.

Filter leaf tests conducted on corn stover utilized the Eimco

Table 15. pH Depression as a Function of $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$ Addition

Initial pH	gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$	Final pH
7.6	0.467	6.9
7.6	0.667	6.8
7.6	1.0	6.7
7.7	4.67	4.2
7.7	6.67	2.8
7.4	8.67	2.6
7.4	10.0	2.5
7.4	11.67	2.4

NY-415 medium. This selection was made for a number of reasons. Firstly, this medium can produce filter cakes without the aid of any chemical conditioning. Secondly, plugging does not appear to be a problem for this filter leaf. In addition, the mass of solids captured per cycle is large. This implies that high solids loading rates will be possible.

The first tests run on the fermented corn stover did not involve chemical conditioning. The effects of varying form and dry time for this experimental mode were studied. Figure 8 shows that drier cakes were produced with longer form and dry times. The driest cake produced without chemical conditioning had a solids content of 20 percent. Increasing the form time beyond the range studied does not appear to substantially increase the cake dryness. A limiting factor in this experimental procedure is the weight of filter cake that can be suspended from the filter leaf upon withdrawal from the slurry. The filter cakes produced in the above tests were in excess of one inch thick. When a cake of this nature is removed from the slurry, it is saturated with fluid and is heavy. When using form times

Figure 8. The Effects of Form Time and Dry Time on the Filter Performance

longer than 30 seconds, the cake usually falls from the filter leaf as it is withdrawn from the slurry. Increasing the drying time beyond 135 seconds increased the filter cake solids due to air drying of the cake. Since the cake releases water very quickly, thus allowing the movement of large volumes of air through the filter cake, extended dry times could tend to be expensive.

The limiting factor in attaining dry cakes with no chemical conditioning was the amount of water present in the large particles (those retained on a 10 mesh sieve). These particles had phenomenally large swelling properties and actually grew to twice their normal size upon wetting. Since the fermented corn stover slurry had two distinct particle size distributions, it is likely that the vast majority of the particles retained by the filter leaf, when no chemical conditioners were used, was in the 10 mesh size category. Therefore, even if the interparticle spaces were completely drained of liquid, a large amount of water is still absorbed by the particles.

Figure 8 also shows the response of the filtrate solids concentration to the test conditions. There was slight variation in the filtrate suspended solids concentration with respect to the form time. Form time also appeared to slightly affect the filtrate dissolved solids concentration which was not a function of filtration. Therefore, it is highly probable that this minor variation actually resulted from a difference in sludge characteristics related to a time lag between running the two series of tests. Also, there was only a slight increase in the filtrate solids concentration as the dry time increased. This could be due to the high permeability of the filter cake with respect to the smaller suspended solids and liquid fraction.

Therefore, the residual amount of small particles to be drained during the final stages of drying was low.

Figure 8 also depicts trends which are peculiar to the filterability of fluid and of small particles through this filter cake and medium. A higher solids capture might be expected to occur at longer form times. This was not the case in this instance. The reason for the difference is that as the medium emerges from the slurry tank, the mass of particles caught on the filter medium and removed from the slurry tank was limited. However, there was no limit to the amount of fluid and small particles which pass through the porous filter cake and filter media. If in the extreme case, the filter apparatus was held in the slurry tank long enough, all of the fluid would be removed. But when lifting the filter apparatus from the beaker, only a portion of the solids initially retained by the filter leaf remained attached to the medium. The remainder of the particles fell back into the slurry beaker. The extreme case, where the 3 liter slurry beaker is dewatered, occurs when using a form time of 45 seconds.

In an attempt to achieve higher cake solids, chemical conditioning with ferric chloride was investigated. For these tests and those to follow, standard form and dry times of 30 seconds and 120 seconds, respectively, were used in order to facilitate data comparison. The results are depicted as the solid symbols in Figure 9. As the dosage of ferric chloride increased, more of the smaller particles were retained by the filter cake. In the case of corn stover, the smaller particles contain less moisture than the larger particles. Therefore, the dryness of the filter cake increased with the dosage of ferric chloride. However, addition of ferric chloride incorporated a gelatinous nature into the filter

Figure 9. The Effect of FeCl_3 Conditioning on the Filter Performance

cake. For this reason, the increase in cake solids was small with respect to the massive amounts of ferric chloride added.

Examination of Figure 9 reveals that the fine suspended material was indeed being coagulated and incorporated into the filter cake. The increase in the dissolved solids concentration reflected the addition of ferric chloride to the slurry.

Figure 9 also shows that approximately 90 percent of the suspended solids can be captured with the use of ferric chloride as a conditioner on the slurries that had not received the thermochemical pretreatment. This implies that the conditioner was fulfilling its role with regard to destabilizing the colloidal system and enhancing particle capture. This information supports the theory that achievement of drier filter cakes was being hindered by the water bound in the large particles and the water associated with the ferric ion.

Similar tests were run on a slurry which had undergone caustic pretreatment prior to digestion. This pretreatment consisted of dosing the corn slurry with NaOH (3 percent NaOH) and cooking it at a pressure of 10 psi for four hours. The caustic treatment generally broke down the structure of the corn particles making them more supple. The results of the vacuum filtration studies on this slurry are shown in Figure 9.

The slurry subjected to caustic treatment prior to digestion formed drier filter cakes than the slurry not subjected to this treatment. As a result of this pretreatment, a higher percentage of the cellulose was converted to gas. Cellulose has an affinity for water. Consequently, a reduction in the quantity of cellulose will result in less retention of internal moisture by the particles.

Figure 9 supports this principle. The residual amount of small suspended solids not incorporated into the filter cake was larger for the caustic treated slurry. So even though less fines were in the cake, it was drier. This implies that there was less moisture retention by the large particles. The increased filtrate suspended solids concentration for the caustic treated slurry was probably due to the higher alkalinity of the caustic slurry. In this case, more ferric chloride was required for pH adjustment than when no caustic pretreatment was used. The lower solids capture merely reflect the higher concentration of filtrate suspended solids encountered with caustic pretreatment.

The filtrate suspended solids obtained with stover slurry that was not subjected to the caustic pretreatment decreased with iron dosages up to about 5 g/l. Additional iron dosage appeared to marginally increase the cake solids. This additional cake solids may have been the result of more iron precipitate in the cake or because of the lower pH resulting from the higher iron dosage. To determine the extent of this dilemma, pH adjustment with acid prior to ferric chloride addition was investigated. It was hoped that smaller iron dosages could be used once an optimum pH was attained through the addition of the acid. For these experiments concentrated sulfuric acid (18 molar) was used in order to avoid dilution of the test slurry. The iron dosage used in this test was 1.67 gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$. Table 16 shows the extent of pH adjustment through the addition of sulfuric acid while the experimental results are shown in Figure 10.

As illustrated in Figure 10, the depression of the pH by the addition of acid did not affect the dryness of the filter cake. The filtrate suspended solids concentration illustrates that an optimum pH for particle destabilization did exist. This optimum pH was approximately

Figure 10. The Effect of pH on Filter Performance

Table 16. pH Depression Through the Addition of Sulfuric Acid and Ferric Chloride

Initial pH	ml* 18 Molar H_2SO_4 Added	Intermediate pH	Final pH After Ferric Salt Addition
8.0	1.0	6.1	5.3
7.8	1.4	5.5	4.0
7.8	1.7	4.6	3.5
8.0	2.2	3.4	2.9
7.8	2.9	2.4	2.4

*1.5 liter slurry test volume

3.5. The cake solids appeared to be a function of iron dosage rather than pH. An increase in solids capture did not influence the cake solids. The optimum pH for solids capture appeared to be about 3.0.

The use of gravity thickening was examined as a pretreatment step to vacuum filtration in an attempt to produce drier filter cakes. This test was carried out by adding ferric salts to a known volume of digester effluent. The mixture was placed in a one liter graduated cylinder. The cylinder was equipped with a slow speed mixer (~1 rpm). The purpose of the slowly rotating device was to eliminate the wall effects associated with the graduated cylinder without disturbing the floc particles.

The treated slurry exhibited zone settling properties. The volume of fluid beneath the clear liquid interface was recorded as a function of time. A value termed percent settling, which is the volume of clear liquid divided by the total volume of fluid, is graphed as a function of time in Figure 13 for varying iron dosages.

As Figure 11 shows for a substantial dosage, 4.67 gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$, less than a 20 percent settling of the interface occurred within 4 hours of the start of the test. For a massive dosage, 30 gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$, there is only a 40 percent settling of the interface in 4 hours time. For thickening to be economical as a pretreatment scheme, one would desire a 50 percent settling of the interface in 2 to 3 hours time with a much smaller chemical addition than those shown in Figure 11. For this reason, the thickening procedure described does not appear to be practical.

Anionic polymers were also used as sludge conditioners. As Figure 12 illustrates, Dow Purifloc A-22 can produce drier cakes than A-23. In addition, it appears that no benefits were accrued by using large polymer dosages. In fact, overdosing the slurry appeared to occur when using A-23 in concentrations greater than 400 mg/l. The filtrate suspended solids concentration was not significantly altered with polymer dosage. An exception appeared to be apparent overdosing which occurs when using A-23 in dosages greater than 400 mg/l. It is important to note that the filtrate suspended solids were high in this battery of tests. In fact, they were twice as high as that which would be produced without any conditioner addition.

The percent solids capture when using A-22 did not vary appreciably with polymer dosage. This is not surprising since the other experimental variables did not vary with polymer dosage for this particular polymer. When using polymer A-23 the solids capture did show some response to polymer dosage. However, there was one data point which appears to be out of proportion with the rest of the data. No explanation other than experimental error can be offered at this time for that data point.

Figure 11. The Effect of Iron Dosage on Slurry Settling Rates

Figure 12. The Effect of Anion Polymer on the Filter Performance

The data obtained when using anionic polymer were very peculiar. The concentration of filtrate suspended solids was higher than when using no coagulant aids. In many cases the dosage of polymer had no effect on experimental results. Yet, the cake percent solids and percent solids capture were appreciably higher than when using no chemical conditioning.

An experimental observation might help explain this situation. When using anionic polymers the thickness of the filter cake was 25 percent to 50 percent less than when no conditioners were employed. Therefore, it could be possible that the filter cakes formed using anionic polymers were easier to air dry since there was a smaller volume of solids to dry. In addition, the volume of filtrate was much smaller when using the anionic polymer which would tend to increase the percent solids capture, overshadowing the increased filtrate percent suspended solids which occurred during these tests.

Another investigation was made using two cationic polymers, Dow Purifloc C-31 and C-43. Figure 13 shows that C-31 did not produce significantly drier cakes than any of the other conditioners that were investigated. However, C-43 did produce filter cakes of appreciable dryness. The filtrate suspended solids concentrations for the two polymers were similar within the bounds of experimental error. This would imply the same degree of particle destabilization and incorporation into the filter cake. Thus, one may hypothesize that the difference in the cake moisture content was due to mode of particle destabilization and the structure of the cakes attained.

Increasing the polymer dosage of C-43 increased the percentage of mass retained by the filter leaf. Experimental records show that as the dosages increased, the volume of filtrate decreased. This caused the

Figure 13. The Effect of Cationic Polymer on Filter Performance

above trend to develop. The converse is true for the case of the cationic polymer C-31. As the polymer dosage increased, so did the volume of fluid which passed through the filter cake and medium. The result of this event was a decreased percent solids capture as polymer dosages increased.

Additional experimentation using increased form pressures was conducted with the hope that some of the moisture residing inside of the corn particles would be removed. Two test series were run. One used no chemical conditioning while the other used a dosage of 6.67 gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$. The non-zero dosage was chosen with the help of Figure 9. Dosages higher than 6.67 gm/l $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$ had no appreciable effects on cake dryness. The experimental results are depicted in Figure 14. The cake solids increased a few percent as the form pressure increased. Of course, the chemically treated slurry produced a drier cake. It is important to realize that with this experimental set-up, an increase in form pressure implies an increase in dry pressure also.

The filtrate suspended solids concentration was lower for the chemically conditioned sludge. This is a logical development. The interesting point is that the filtrate suspended solids concentration increased with increasing form pressure while the converse was true for the unconditioned sludge. The reason for this development must lie in the structural integrity of those particles chemically flocculated and how they withstand the stress imposed by increasing form and dry pressures. There was only a slight difference between the solids capture with respect to the two slurries investigated.

The solids loading rate is a very important design parameter with respect to equipment selection and sizing. This parameter describes the weight of solids (on a dry basis) which is retained per square foot

Figure 14. The Effect of Pressure Differential on Filter Performance

of filter media per hour. Reviewing the objectives of this dewatering process, it is appropriate to cite the solids loading rate for a number of pertinent system conditions such as: obtaining the driest filter cake, minimum chemical addition, lowest effluent solids concentration, and highest solids capture. The results are listed in Table 17.

The parameters listed in Table 17-A pertaining to production of the driest filter cake relate to a polymer dosage of 160 mg/L. Review of Figure 13 shows that a slightly drier cake can be formed with a dosage of 200 mg/L. The difference is insignificant when considering experimental error. Therefore, the lower dosage, namely 160 mg/L, was cited in Table 17-A.

Values given in Table 17-B relate to a form time of 30 seconds and a dry time of 135 seconds since these conditions yield the driest filter cake when no chemicals are used. Table 17-C relates to test conditions which produce the lowest filtrate total solids concentration. There were tests which yielded lower suspended solids concentration. However, the filtrate dissolved solids concentration were excessive due to large chemical additions.

When one considers the amount of nutrients that must be added to the corn stover slurry in order to maintain the digestion process, re-use of the effluent liquid fraction becomes an interesting prospect. The influent digester slurry is generated by incorporating the freshly ground corn substrate into the reclaimed liquor. Not only does this effluent utilization method reduce the cost of supplemental nutrient addition, but also eliminates the transportation costs which are generally associated with the alternative utilization plans.

Land application of liquid fraction is also possible. However, the available amount of nutrients is minimal when considering agricultural

Table 17. Solids Loading Rates

(A) DRIEST FILTER CAKE

Cake Solids = 28.7%	Form Time = 30 seconds
Filtrate Suspended Solids = 1.0%	Dry Time = 120 seconds
Filtrate Dissolved Solids = 0.33%	Conditioner: Dow Purifloc C-43
Solids Capture = 65%	Dosage: 160 mg/l

$$\text{Solids Loading Rate} = 1.64 \text{ lb/ft}^2 - \text{hr}$$

(B) MINIMUM CHEMICAL DOSAGE

Cake Solids = 20.2%	Form Time = 30 seconds
Filtrate Suspended Solids = 0.81%	Dry Time = 135 seconds
Filtrate Dissolved Solids = 0.47%	Conditioner: None
Solids Capture = 60.1%	Dosage: None

$$\text{Solids Loading Rate} = 5.67 \text{ lb/ft}^2 - \text{hr}$$

(C) LOWEST FILTRATE SOLIDS CONCENTRATION

Cake Solids = 20.2%	Form Time = 30 seconds
Filtrate Suspended Solids = 0.5%	Dry Time = 120 seconds
Filtrate Dissolved Solids = 0.54%	Conditioner: $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$
Solids Capture = 87.4%	Dosage: 4.67 g/l

$$\text{Solids Loading Rate} = 3.79 \text{ lb/ft}^2 - \text{hr}$$

(D) HIGHEST SOLIDS CAPTURE

Cake Solids = 21.9%	Form Time = 30 seconds
Filtrate Suspended Solids = 0.41%	Dry Time = 120 seconds
Filtrate Dissolved Solids = 0.79%	Conditioner: $\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$
Solids Capture = 90.6%	Dosage: 10.0 g/l

$$\text{Solids Loading Rate} = 5.31 \text{ lb/ft}^2 - \text{hr}$$

requirements. This plan would also entail added costs due to transportation.

A number of plans can be enumerated with regard to the utilization of the digester effluent solid fraction. If the digestion units are located close to feedlots, the solid fraction can serve as a matrixing substrate for various feed nutrients. If such an option is not available, the solid fraction can be applied to the farmland. The solids would act as a soil conditioner in addition to returning minerals to the soil.

Power generation through incineration of the solids fraction does not appear to be a favorable disposal option. A 30 percent filter cake solids contents was never achieved through vacuum filtration. Therefore, auxiliary fuel would have to be employed to maintain combustion. Alternate dewatering systems such as filter belt presses may make this a viable option for a portion of the residue.

Reviewing the corn stover digester effluent vacuum filtration data, filtration without the use of chemical conditioners appears to be the most reasonable course to follow. Increased cake dryness is marginal even under massive dosages of conditioners. In addition, a large percent solids capture or low filtrate solids concentration is not necessary in light of the properties of the substrate or the effluent utilization plans outlined above.

In some instances, when transportation costs are high, the production of drier filter cakes may be warranted. This can be done with increased form pressure or addition of chemical conditioners. The conditioners of choice would be FeCl_3 or Dow Purifloc C-43, or other appropriate polymers. The extent of the form pressure increase and/or chemical conditioning is governed by the savings accrued in transportation costs.

The above disposal techniques should by no means be viewed as the only solutions available. Other solids separation techniques such as centrifugations, filter presses, and belt presses should also be evaluated since the uniqueness of each process has a bearing on the cost effectiveness of the various effluent utilization schemes.

Centrifuge Studies

Dewatering of the reactor effluents by centrifugation is an alternate to dewatering by filtration. A fletcher Mark III solid bowl basket centrifuge with a variable speed drive was used to determine the dewatering characteristics of the fermented corn slurry. The feed slurry is continuously pumped into the center feed tube from which it is discharged at the bottom of the bowl. The centrate flows vertically and is discharged at the top of the bowl. This unit does not provide for a continuous cake discharge. When the cake reaches a predetermined level, the feed is discontinued.

Reactor effluent was pumped to the centrifuge at a constant rate using a Moyno pump. Most of the runs were made with a feed rate of 8 lpm. The centrifuge was brought up to speed and the pump started. This was continued until approximately 40 liters had been pumped through the centrifuge. Since the pool volume of this centrifuge was 10.2 liters; the feed volume allowed for four volume displacements. Before the feed pump was stopped, a sample of the centrate was obtained. As soon as the feed pump was stopped, all of the centrate in the bowl was removed with a skimmer pipe. Only the cake remained in the bowl. This cake was removed manually and a sample obtained.

The centrifugal forces generated by the machine are a function

of the rotational speed and the bowl diameter. The bowl diameter of this machine was 35.6 cm (14 inches) and the maximum speed was 3400 rpm. Under these conditions, the maximum centrifugal force was $2300 \times g$.

The dewatering characteristics of the fermented corn slurries generated under various operating conditions were evaluated using the basket solid-bowl centrifuge. The results of an initial run are shown in Figure 15. Reactor 1 was receiving corn stover slurry that had been thermochemically pretreated with 10 percent by weight (dry solids) of lime. The feed to Reactor 3 was untreated corn stover. Both reactors were operated at 58°C and a 14.7 day retention time. Since the lime pretreatment did not improve the stover biodegradability, the conversion efficiency of both reactors was approximately the same.

The maximum cake solids were obtained at 2400 ($1145 \times g$) and 2600 rpm ($1343 \times g$), the maximum speeds used on the samples from Reactors 1 and 3, respectively. It would appear that higher speeds would produce a drier cake. The cake produced from the lime pretreated slurry had a solids content of almost 38 percent as contrasted to the 30.5 percent obtained with control reactor. The added lime clearly produced a more easily dewatered sludge.

There was a marked difference in the solids content of the centrate produced in these tests. This may be due in part to the high solids content (5.3 percent) of Reactor 1 compared to the control reactor (3.44 percent). The total solids in the centrate from Reactor 1 were almost double that from Reactor 3. The higher soluble solids in Reactor 1 did not account for this difference. Suspended solids were the primary difference. This was reflected in the suspended solids capture. For Reactor 1, the solids capture increased from 67.1 percent at 1000 rpm

Figure 15. Fermenter Slurry Dewatering by Centrifugation

to 74.4 percent at 2400 rpm. The solids capture from Reactor 3 was somewhat better, ranging from 70.5 percent at 1000 rpm to 78.5 percent at 2600 rpm.

The relatively flat nature of the effluent solids curve shows that there is little possibility of increasing solids capture with increased centrifugal force. Chemical conditioning of the slurries would be necessary to increase the solids capture. As the data from the filtration studies show, significant chemical dosages were required for effective capture of these small particles. These chemicals add significantly to the cost of the slurry dewatering. Since centrate recycle is highly probable, high solids capture is less important. These solids can be allowed to accumulate in the system, reaching some equilibrium value.

The pretreatment of these fibers with sodium hydroxide changed the solids characteristics to a degree. As discussed earlier, the caustic pretreatment followed by fermentation resulted in a higher percentage of smaller size particles. The higher the caustic dosage and the longer the fermenter retention time, the greater the effect. Figure 16 presents the results of two centrifuge runs each on two reactors receiving stover that had been treated with 3 percent NaOH. Reactor 1 was operated at 14.7 day retention time while Reactor 2 was operated at 7.8 days. The maximum cake solids for Reactor 1 effluent was 28 percent at 3000 rpm ($1788 \times g$). Reactor 2 effluent yielded a slightly higher cake solids of 29 percent.

The solids content of the centrate was significantly higher than the control (Reactor 3) in Figure 15. This centrate had a total solids content of about 1.1 percent, of which 0.4 percent was soluble.

Figure 16. Dewatering Characteristics of Thermochemical Pretreated Slurry

The centrate from Reactor 2 effluent contained approximately 1.3 percent total solids. Analysis of this slurry for filtration studies yielded a soluble solids of 0.65 percent. Therefore, the total solids increase in this slurry was due to the increase in soluble solids. Reactor 1 effluent produced a centrate with approximately 1.6 percent total solids. With a soluble solids of 0.65 percent, the increased solids would be, in part, a result of a higher concentration of suspended solids.

Solids capture was not high with these slurries, as was the case with the effluent from the control reactor. Reactor 2, operating at the shorter retention time, exhibited a solids capture ranging from 78.8 percent at 1500 rpm to 87.0 percent at 3000 rpm. This was somewhat better than for the control. This may have been a result of a higher feed slurry solids to the centrifuge from Reactor 2. The solids capture for Reactor 1 was comparable to that obtained from the control, ranging from 72.0 percent at 1500 rpm to 78.7 percent at 2500 rpm. The centrate suspended solids were substantially greater than the control, but the higher slurry solids in the centrifuge feed resulted in a large solids capture percentage.

The data in Figure 17 present the results of two centrifuge runs on each reactor receiving pretreated corn stover with a NaOH dosage of 5 percent. Reactor 1 was operated at a 14.7 day retention time while Reactor 2 was operated at 7.8 day retention time. Both slurries produced a dry cake, with solids contents in the 33 to 36 percent range. The effluent solids were high, between 1.6 to 1.7 percent total solids. Data on soluble solids were not obtained; therefore, solids capture cannot be calculated. However, since both of these slurries had total solids less than 4.0 percent, the suspended solids capture will not be high. If one assumes a soluble solids of 0.9 percent, the solids capture for Reactor 1

Figure 17. Dewatering Characteristics of Thermochemical Pretreated Slurry

at 3000 rpm would be 77.5 percent. Efficient solids capture will require chemical addition.

These data show that the centrifuge can be fairly successful in removing particles over a wide size range. The particle size distribution for Reactor 1 (see Table 12) shows that only 37.8 percent of the total solids were greater than 74 μm size. Clearly the centrifuge has the ability to remove a significant fraction of these particles that pass the 200 mesh screen.

PROCESS EVALUATION

In order to obtain some insight into the economics and energy efficiency of a corn stover fermentation system, a simple simulation of the system was developed. The processing system is divided into several subsystems as follows:

- Preprocessing
- Fermentation
- Slurry Dewatering
- Gas Cleanup
- Boiler System

Since the stover is assumed to be chopped for increased density during collection and transport, size reduction was not considered part of the preprocessing subsystem. Without thermochemical pretreatment this subsystem contains only a premix tank that is used to slurry the dry stover with water. Steam is injected into this tank to increase the temperature of the feed slurry. With thermochemical pretreatment, pressure reactors are installed prior to the premix tank. Treated stover is discharged to the premix tank where the solids content is adjusted to the desired level for feed to the fermentation tanks.

Fermentation is conducted in mixed single stage reactors. Additional steam is injected directly into these tanks to maintain the thermophilic fermentation temperature of $59 \pm 1^\circ\text{C}$. Since the feed slurry is heated to a few degrees above the fermentation temperature, limited steam is injected into the fermentors.

Heat recovery is accomplished by one of two methods. If thermochemical pretreatment is not used, centrate from the centrifuge is recycled

for slurry water for the stover. This water retains most of its temperature thereby significantly reducing the steam required to heat the feed slurry. When thermochemical pretreatment is used, recycle is not possible. In this case, slurry water is passed through a heat exchanger on the fermentor effluent. This captures a significant portion of the sensible heat in the fermented slurry.

Slurry dewatering is accomplished by centrifugation in the simulator. Alternate dewatering techniques are available that may be less costly than centrifugation. Depending upon the technique employed for return of the residue to the land, dewatering may or may not be practiced when thermochemical pretreatment is employed. Without this pretreatment, it may be economically attractive to dewater the slurry in order to conserve the heat, alkalinity, and nitrogen and phosphorus as well as the recycled water, and also to reduce disposal costs.

A boiler system is necessary in order to provide the heat energy required. There is no fuel cost assigned to the boiler as the gas from the fermentor is used in the boiler and the product gas yield reduced accordingly. Gas cleanup via the monoethanolamine system has been included, but it is highly doubtful if it is necessary or economically feasible to produce pipeline quality gas.

A very simple cost calculation is used. Capital costs are not total capital, but represent only the equipment acquisition and installation costs and the construction costs associated with structures. These costs do not include such costs as engineering, land, interest during construction, etc. The hourly costs represent only the amortization of the above capital (10 percent interest over 20 years) plus process operating and maintenance costs. No attempt has been made to use the utility cost accounting

system since these simulations are being used for comparison purposes only.

If one wishes to place a price on the gas, substrate acquisition costs (\$20/tonne dry solids) of \$642/hr plus additional costs for estimating total capital requirements, plus \$3/tonne of wet residue disposal costs must be added to the cost developed in the computer program. This cost would relate to public financing without taxes, return on investment, etc.

Simulation Results

The response of the system is shown in Table 18. Four runs were made to illustrate the process performance under various operating conditions. Costs are expressed as 1979 dollars and the product gas was priced at \$1.90/GJ ($\$2/10^6$ BTU) in one case and for comparative purposes at \$7.60/GJ. The energy balance is based on an input energy of the stover of 15 GJ per tonne of stover processed. The purchased electrical power, produced at a 30 percent efficiency, is added to this energy content of the stover to calculate the total energy input. A more detailed explanation of the simulation model is contained in Appendix C.

Run 1 represents a base line system receiving corn stover at the rate of 1000 tonnes per day. The reactor slurry solids are fed at 8 percent due to mixing problems and costs associated with higher solids concentration. These data clearly show that corn stover with a biodegradability factor of only 40 percent is not economically feasible. Gas at any price will not pay for processing stover simply because the energy content of the product gas is only slightly greater than the process heat requirements. Addition of the electrical power production causes the purchased energy to exceed the product energy.

Table 18. Results of Simulation Runs - 1000 Tonnes/Day

Processing Conditions	Run 1	Run 2	Run 3	Run 4
Recycle - %	50	50	50	0
Biodegradability - %	40	90	90	77
Feed slurry solids - %	8.9	8.9	16.4	12.0
Digester solids - %	8.0	4.2	8.0	7.5
θ - days	7.0	7.0	7.0	7.0
<u>Fermentation System</u>				
Volume - 1000 m ³	60.6	60.6	27.5	45.0
Cost - \$10 ⁶	6.87	6.87	3.12	5.12
Heat req'd - GJ/hr	51.7	52.6	30.7	49.4
Chemical costs - \$/hr	-	-	-	167.50
<u>Centrifuge System</u>				
Machines	20	10	10	13
Cost - \$10 ⁶	6.89	3.45	3.45	4.49
<u>Economic Summary</u>				
Equip. & instal. cost - \$10 ⁶	14.14	10.72	6.97	11.69
Total cost - \$/hr	420	268	205	471
Gas value (\$1.90/GJ) - \$/hr	5.6	417	475	271
"Net" income - \$/hr	(415)	149	270	(220)
Gas value (\$7.60/GJ) - \$/hr	22.4	1668	1900	1084
"Net" income - \$/hr	(398)	1400	1695	593
<u>Energy Balance</u>				
Product energy - GJ/hr	2.9	220	250	143
Purchased power - GJ/hr	16	8	8	11
% Input energy recovery*	0.6	45	51	29

*Based on stover feed rate of 32.12 tonnes/hr at 482 GJ/hr energy content.

Run 2 and Run 3 were designed to illustrate the extreme importance of the biodegradability factor. If the biodegradability of the substrate volatile solids increases to 90 percent while the remaining parameters are held constant, one can see a major economic gain. First, the product gas quantity greatly increases as a result of a much larger quantity of solids being converted to gas. Second, the dewatering cost is reduced by a factor of two since the number of centrifuges is reduced from 20 to 10. Obviously, this results in a concomitant reduction in capital, operating and maintenance costs. Even though the water quantity is essentially constant between these two runs, the substantial reduction in solids flow rate for Run 2 results in a requirement for 10 machines since the solids flow rate is the limiting factor for the centrifuge operation.

The biodegradability factor alone does not improve the economics sufficiently to consider the system feasible. A \$20/tonne for substrate acquisition cost adds \$642/hr to the total cost. At the higher gas price, a "net" income of \$1400/hr will begin to approach feasibility when all costs are included. The increased biodegradability is important in reactor sizing since the biodegradability controls the relationship between feed slurry solids and fermentation slurry solids concentration. Based on the experiences with mixing of fermenters processing urban refuse, a slurry concentration of 8 percent in the fermenter has been used. Data are presently unavailable on the mixing properties of slurries from various crop residues.

In Run 1, the feed solids concentration was fixed at 8.9 percent in order to satisfy the 8 percent limit in the fermenter. For comparison, Run 2 was made with a limit of 8.9 percent on the feed solids concentration. Because of the high biodegradability factor, the fermenter solids were

reduced to less than one-half of the feed solids. In order to take full advantage of the high conversion efficiency, a higher feed solids can be employed such that the reactor slurry solids will be 8 percent. The effect of the higher feed solids is dramatic. The fermenter volume is reduced to less than one-half of Run 1. The heat requirements for fermenter heating are reduced from 56.6 to 30.7 GJ/hr resulting in a net increase in product gas energy of 30 GJ/hr, using a boiler efficiency of 0.85. As a result of the increase in product gas and the reduced capital and operating costs, a "net" income of \$1695/hr is realized. The economic potential of this system will be discussed more later.

Increased biodegradability of corn stover can be achieved through a thermochemical pretreatment process as discussed previously. Run 4 shows the results of this analysis with an increase in the biodegradable volatile solids to 77 percent. The advantages of Runs 2 and 3 are present in Run 4. Unfortunately, a significant chemical cost of \$167.50 (\$5.20/tonne of dry stover or \$1.17/GJ of energy produced) is added to the processing costs.

A more detailed economic analysis of Runs 3 and 4 have been made to estimate the likely cost of gas produced by these systems and the importance of improving the biodegradability factor of the substrate. This is presented in Table 19. Capital is available at 10 percent and public financing is used, i.e., no taxes, profits, etc. Clearly, the thermochemical pretreatment changes the process from a complete failure to one that can produce gas for perhaps \$10/GJ. The importance of this pretreatment cannot be over-emphasized. If the biodegradability of the corn stover volatile solids could be increased to 90 percent, up to \$6/GJ could be spent on this pretreatment step and still be equal to Run 4. Essentially, significant savings can be

Table 19. Detailed Economic Analysis for Runs 3 and 4

	<u>Run 3</u>	<u>Run 4</u>
Plant Costs - $\$10^6$		
Equipment and installation costs	6.57	11.69
Support facilities - 5%	<u>0.33</u>	<u>0.58</u>
Total Capital Investment	6.90	12.23
Contractor overhead and profit - 10%	0.69	1.22
Engineering and design - 10%	<u>0.69</u>	<u>1.22</u>
Subtotal Plant Investment	8.28	14.67
Project contingency - 15%	<u>1.24</u>	<u>2.20</u>
Total Plant Investment	9.52	16.87
Interest during construction - 10%	0.95	1.69
Working capital - 2%	<u>0.19</u>	<u>0.34</u>
Total Capital Requirements	10.66	18.90
Amortized Capital Costs - \$/hr	157	278
Operating and Maintenance Costs - \$/hr	81	146
Substrate Acquisition Costs - \$/hr	642	642
Chemical Costs - \$/hr	-	167
Residue Disposal Costs - \$/hr	<u>122</u>	<u>163</u>
Total Cost - \$/hr	1002	1396
Product Gas - GJ/hr	250	143
Gas Cost - \$/GJ	4.00	9.76

expected if techniques can be developed to approach 100 percent conversion. The reduction in fermenter cost and fermenter slurry handling costs will be very great.

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APPENDIX A

The tables in Appendix A present the actual data obtained from the various reactors under the test conditions being employed. The date is designated as the year and the day of the year. For example 78014 is January 14, 1978. The time is listed as the hour of the day. Also, any missing data are recorded as -1. This technique allows the use of zero when a data point is zero. The gas meter readings are the actual meter readings in liters. The difference between two days readings represents the gas produced during that time period. The alkalinity, solids and volatile acids are all expressed as mg/% of the appropriate constituent.

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EPP. T.S.	EPP. V.S.	EPP. V.A.	CH4 %
78014										
10.45	1431.00	57.00	7.20	-1.00	34052.00	24772.00	-1.00	-1.00	-1.00	-1.00
78015										
14.30	1767.00	51.00	7.15	-1.00	29800.00	21109.00	-1.00	-1.00	-1.00	-1.00
78016										
11.10	2029.00	48.00	7.10	1425.00	39608.00	27686.00	21644.00	12991.00	72.00	58.30
78017										
8.25	2300.00	50.00	7.20	1472.00	35786.00	25785.00	21759.00	13006.00	96.00	58.80
78018										
8.20	2581.00	55.00	7.20	1330.00	36726.00	26311.00	23208.00	14658.00	96.00	54.90
78019										
8.10	2910.00	55.00	7.20	1330.00	37665.00	26837.00	22797.00	14335.00	72.00	55.00
78020										
8.25	220.00	51.00	7.20	1330.00	29271.00	19514.00	23308.00	14755.00	96.00	54.80
78021										
13.00	610.00	52.00	7.20	-1.00	32368.00	21000.00	-1.00	-1.00	-1.00	-1.00
78022										
13.50	907.00	53.00	7.00	-1.00	30924.00	21209.00	-1.00	-1.00	-1.00	-1.00
78023										
8.15	1175.00	48.00	7.20	1330.00	32716.00	22774.00	23219.00	14398.00	168.00	52.80
78024										
8.15	1486.00	62.00	7.20	1188.00	33434.00	23205.00	22545.00	13922.00	120.00	52.70
78025										
8.25	1856.00	53.00	7.20	1188.00	32842.00	22789.00	22242.00	13709.00	96.00	53.10
78026										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	55.10
78027										
12.10	2388.00	60.00	7.10	1235.00	32193.00	22215.00	22612.00	13709.00	144.00	55.10
78028										
10.30	2595.00	54.00	6.90	-1.00	33490.00	23259.00	-1.00	-1.00	-1.00	-1.00
78029										
12.40	2902.00	54.00	6.90	-1.00	33410.00	23285.00	-1.00	-1.00	-1.00	-1.00
78030										
8.20	122.00	57.00	7.20	1188.00	29128.00	19407.00	23974.00	14698.00	120.00	52.50
78031										
8.10	396.00	52.00	7.20	1235.00	29648.00	19523.00	22927.00	14026.00	72.00	54.70
78032										
8.10	659.00	54.00	7.20	1235.00	28502.00	18862.00	22393.00	13713.00	72.00	55.70
78033										
8.15	918.00	54.00	7.20	1140.00	30935.00	20578.00	23027.00	14007.00	72.00	55.20
78034										
8.15	1170.00	62.00	-1.00	1235.00	29089.00	19682.00	23253.00	14302.00	48.00	55.00
78035										
9.35	1459.00	54.00	7.00	-1.00	28454.00	18405.00	-1.00	-1.00	-1.00	-1.00
78036										
9.20	1758.00	50.00	7.20	-1.00	28417.00	19175.00	-1.00	-1.00	-1.00	-1.00
78037										
8.20	1993.00	74.00	7.20	1188.00	31123.00	21039.00	23513.00	14409.00	96.00	52.20
78038										
8.20	2234.00	49.00	7.20	1188.00	34081.00	24435.00	22557.00	13530.00	120.00	52.70

Table A-1a. CORN STOWLER REACTOR NO.1 JANUARY 78

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	PE20 T.S.	PE20 V.S.	2PP. T.S.	2PP. V.S.	EPP. V.A.	CH4 %
78039										
8.10	2355.00	66.00	7.20	1235.00	33754.00	21957.00	22554.00	13652.00	120.00	59.18
78040										
8.10	2537.00	53.00	7.20	1188.00	32012.00	22173.00	22948.00	13900.00	96.00	56.90
78041										
8.15	2917.00	52.00	7.10	1141.00	30897.00	21972.00	22839.00	14231.00	144.00	51.50
78042										
11.30	333.00	51.00	7.00	-1.00	37354.00	25715.00	-1.00	-1.00	-1.00	-1.00
78043										
13.00	687.00	56.00	7.20	-1.00	35198.00	23566.00	-1.00	-1.00	-1.00	-1.00
78044										
9.10	971.00	59.00	7.10	1235.00	40964.00	27031.00	24041.00	14717.00	120.00	52.90
78045										
8.25	1355.00	61.00	7.20	1235.00	41238.00	28723.00	24483.00	14668.00	72.00	51.30
78046										
8.15	1655.00	60.00	7.30	1235.00	43488.00	31017.00	24077.00	14548.00	312.00	53.10
78047										
8.15	2049.00	53.00	7.20	1188.00	46177.00	32754.00	25645.00	15959.00	144.00	54.80
78048										
8.15	2320.00	61.00	7.20	1188.00	38109.00	26566.00	25163.00	15158.00	120.00	55.00
78049										
8.40	2602.00	55.00	7.20	-1.00	37440.00	26533.00	-1.00	-1.00	-1.00	-1.00
78050										
7.25	2860.00	54.00	7.20	-1.00	41740.00	30349.00	-1.00	-1.00	-1.00	-1.00
78051										
8.15	169.00	51.00	7.10	1235.00	40394.00	28862.00	24673.00	15027.00	192.00	50.60
78052										
8.15	432.00	50.00	7.20	1235.00	36513.00	25557.00	25265.00	15455.00	168.00	53.20
78053										
8.15	694.00	71.00	7.20	1188.00	41157.00	29090.00	24141.00	14390.00	144.00	54.20
78054										
8.15	991.00	54.00	7.10	1188.00	35117.00	24319.00	24667.00	15046.00	120.00	53.90
78055										
8.20	1266.00	60.00	7.20	1235.00	45202.00	32741.00	25805.00	16088.00	120.00	54.10
78056										
9.50	1551.00	51.00	7.10	-1.00	45500.00	32126.00	-1.00	-1.00	-1.00	-1.00
78057										
9.10	1884.00	47.00	7.10	-1.00	43731.00	30992.00	-1.00	-1.00	-1.00	-1.00
78058										
8.15	2160.00	54.00	7.00	1292.00	42719.00	30340.00	25824.00	15468.00	240.00	51.30
78059										
8.00	2511.00	46.00	7.20	1235.00	40599.00	26187.00	27399.00	16798.00	768.00	49.80
78060										
7.50	2773.00	63.00	7.20	1330.00	39632.00	2717.00	26841.00	15964.00	96.00	53.30
78061										
7.35	44.00	52.00	7.30	1377.00	55434.00	34158.00	2833.00	16977.00	96.00	55.30
78062										
7.50	462.00	50.00	7.20	1377.00	56515.00	39064.00	29178.00	17860.00	72.00	56.40
78063										
11.00	937.00	62.00	7.10	-1.00	60658.00	43475.00	-1.00	-1.00	-1.00	-1.00

Table A-1a. CORN STOVER REACTOR NO. 1 JANUARY 79

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EPP, T.S.	EPP, V.S.	EPP, V.A.	CH4 %
78064 11.30	1429.00	53.00	7.10	-1.00	61222.00	43005.00	-1.00	-1.00	-1.00	-1.00
78065 8.20	1812.00	53.00	7.10	1472.00	53491.00	33040.00	32171.00	18945.00	168.00	52.00
78066 8.05	2236.00	48.00	7.30	1615.00	58816.00	34444.00	33577.00	18310.00	432.00	48.37
78067 11.50	2672.00	51.00	7.40	-1.00	64445.00	43051.00	34764.00	21886.00	-1.00	52.90
78068 8.20	73.00	47.00	7.20	1662.00	60510.00	42780.00	32472.00	19762.00	144.00	56.19
78069 8.20	443.00	60.00	7.30	1710.00	55093.00	32477.00	34236.00	17933.00	168.00	49.00
78070 11.00	840.00	63.00	7.30	-1.00	67391.00	47576.00	-1.00	-1.00	-1.00	-1.00
78071 12.00	1347.00	50.00	7.40	-1.00	63924.00	44519.00	-1.00	-1.00	-1.00	-1.00
78072 8.20	1711.00	57.00	7.40	1710.00	59531.00	39787.00	38124.00	23606.00	192.00	52.40
78073 8.15	2137.00	49.00	7.40	1758.00	58789.00	40497.00	40934.00	25750.00	144.00	52.20
78074 8.15	2527.00	56.00	7.40	1662.00	70052.00	46910.00	39687.00	24529.00	216.00	56.30
78075 8.15	72.00	54.00	7.30	1805.00	71916.00	47654.00	41753.00	26000.00	168.00	57.20
78076 8.20	567.00	50.00	7.30	1795.00	65713.00	35752.00	41002.00	21502.00	216.00	58.60
78077 8.00	1066.00	49.00	7.40	-1.00	62148.00	44300.00	-1.00	-1.00	-1.00	-1.00
78078 9.00	1576.00	51.00	7.20	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78079 8.15	1986.00	61.00	7.30	2024.00	64085.00	42819.00	47169.00	28756.00	360.00	54.00
78080 8.05	2390.00	49.00	7.30	2064.00	87813.00	50791.00	47732.00	29052.00	192.00	53.60
78081 8.25	2939.00	57.00	7.50	2160.00	77445.00	47228.00	48952.00	29094.00	240.00	54.70
78082 8.10	552.00	54.00	7.40	2160.00	80704.00	51456.00	50772.00	30427.00	216.00	56.40
78083 -1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78084 -1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78085 -1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78086 -1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78087 -1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78088 8.15	2628.00	40.00	7.40	2256.00	82677.00	51780.00	54243.00	30767.00	360.00	49.40

Table A-1a. CORN STOVER REACTOR NO. 1 JANUARY 78

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EPP. T.S.	EPP. V.S.	EPP. V.A.	CH4 %	
78089	8.20	285.00	59.00	7.40	2400.00	82967.00	49728.00	55435.00	31693.00	288.00	53.80
78090	8.15	1002.00	60.00	7.40	2352.00	81954.00	54569.00	58656.00	33937.00	288.00	55.50
78091	13.30	1687.00	55.00	7.20	-1.00	87121.00	58802.00	-1.00	-1.00	-1.00	-1.00
78092	13.00	2493.00	53.00	7.40	-1.00	85715.00	57544.00	-1.00	-1.00	-1.00	-1.00
78093	8.15	439.00	57.00	7.50	2324.00	86777.00	57663.00	62150.00	34570.00	336.00	52.60
78094	8.10	1196.00	55.00	7.40	2400.00	68325.00	43433.00	60504.00	33418.00	528.00	55.40
78095	8.15	1836.00	52.00	7.40	2544.00	66234.00	43742.00	59028.00	34266.00	480.00	54.40
78096	8.30	2416.00	55.00	7.30	2592.00	64512.00	44114.00	57771.00	33297.00	312.00	53.40
78097	8.25	2954.00	61.00	7.40	2640.00	61754.00	41309.00	56490.00	32342.00	264.00	55.00
78098	8.25	492.00	50.00	7.30	-1.00	63905.00	36713.00	-1.00	-1.00	-1.00	-1.00
78099	8.15	999.00	54.00	7.40	-1.00	57448.00	38129.00	-1.00	-1.00	-1.00	-1.00
78100	8.30	1526.00	87.00	7.50	2688.00	59108.00	37575.00	54774.00	31638.00	216.00	50.40
78101	8.30	2049.00	104.00	7.40	2784.00	61852.00	39912.00	53581.00	28873.00	240.00	59.20

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78102	8.30	2639.00	98.00	7.40	2640.00	53831.00	39506.00	51892.00	30016.00	216.00	56.30
78103	8.20	159.00	104.00	7.40	2448.00	65664.00	43857.00	49315.00	27061.00	192.00	55.60
78104	8.25	573.00	117.00	7.30	2304.00	53961.00	39803.00	44486.00	26347.00	144.00	53.90
78105	9.00	952.00	105.00	7.30	-1.00	56108.00	36108.00	-1.00	-1.00	-1.00	-1.00
78106	10.00	1332.00	100.00	7.20	-1.00	45056.00	31780.00	-1.00	-1.00	-1.00	-1.00
78107	8.15	1720.00	112.00	7.30	1920.00	46796.00	33338.00	35392.00	20535.00	216.00	50.40
78108	8.10	2140.00	99.00	7.20	1632.00	51104.00	36799.00	33659.00	19784.00	216.00	51.90
78109	8.10	2515.00	106.00	7.20	1584.00	53246.00	37526.00	33605.00	20135.00	192.00	52.60
78110	8.25	16.00	104.00	7.10	1440.00	51331.00	35370.00	34581.00	21279.00	120.00	51.30
78111	8.25	533.00	95.00	7.10	1440.00	59551.00	39802.00	34929.00	21193.00	144.00	51.60
78112	9.45	1135.00	103.00	7.10	-1.00	58504.00	38685.00	-1.00	-1.00	-1.00	-1.00
78113	10.30	1593.00	100.00	7.10	-1.00	57290.00	38441.00	-1.00	-1.00	-1.00	-1.00
78114	8.30	1919.00	123.00	7.10	1200.00	58100.00	39680.00	31697.00	19149.00	264.00	56.60
78115	8.05	2238.00	109.00	7.10	1200.00	55193.00	36744.00	32685.00	19502.00	240.00	53.70
78116	8.10	2667.00	103.00	7.10	1104.00	54959.00	36095.00	34592.00	20757.00	264.00	55.80
78117	8.30	165.00	104.00	7.10	1200.00	56424.00	34391.00	36487.00	22351.00	264.00	55.40
78118	8.15	565.00	98.00	7.20	1104.00	54728.00	29528.00	32245.00	16295.00	216.00	55.60
78119	8.05	884.00	100.00	7.20	-1.00	48277.00	25680.00	-1.00	-1.00	-1.00	-1.00
78120	8.15	1301.00	104.00	7.00	-1.00	62476.00	43037.00	-1.00	-1.00	-1.00	-1.00
78121	8.25	1717.00	123.00	7.30	1008.00	63664.00	41516.00	33064.00	19364.00	312.00	56.40
78122	8.25	2117.00	116.00	7.10	1056.00	58856.00	36547.00	32952.00	18958.00	144.00	55.60
78123	8.20	2595.00	103.00	7.10	960.00	63156.00	42382.00	35011.00	20439.00	240.00	52.80
78124	8.25	25.00	98.00	7.10	1404.00	62230.00	41782.00	36944.00	21712.00	96.00	52.30
78125	8.15	715.00	97.00	7.00	960.00	59199.00	38948.00	38547.00	22817.00	240.00	51.60
78126	9.30	1272.00	100.00	7.10	-1.00	60167.00	38032.00	-1.00	-1.00	-1.00	-1.00

Table A-1b. START 4-12-78 CORN REACTOR #1 100 L/DAY DBT. TIME =7.75 DAYS

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78127	10.30	1899.00	102.00	7.10	-1.00	65248.00	39655.00	-1.00	-1.00	-1.00	-1.00
78128	8.20	2399.00	103.00	7.10	1104.00	62795.00	38353.00	42092.00	25092.00	288.00	50.00
78129	8.12	86.00	110.00	7.00	1104.00	60095.00	35332.00	42137.00	23986.00	360.00	51.50
78130	8.40	740.00	-1.00	7.20	1104.00	65547.00	41202.00	43354.00	25284.00	408.00	52.10
78131	8.10	1302.00	101.00	7.00	1152.00	66448.00	37628.00	45446.00	26353.00	312.00	-1.00
78132	8.15	1854.00	110.00	7.00	1152.00	65991.00	42758.00	45734.00	26524.00	552.00	-1.00
78133	13.25	2486.00	102.00	7.00	-1.00	64775.00	41138.00	-1.00	-1.00	-1.00	-1.00
78134	12.00	2987.00	104.00	7.30	-1.00	63316.00	39865.00	-1.00	-1.00	-1.00	-1.00
78135	8.05	471.00	96.00	7.20	1200.00	61501.00	33077.00	48994.00	25743.00	528.00	-1.00
78136	8.10	1064.00	60.00	7.20	1296.00	114043.00	66719.00	48238.00	27087.00	360.00	-1.00
78137	8.15	1489.00	54.00	7.30	1248.00	102563.00	64221.00	49664.00	28056.00	336.00	54.60
78138	8.20	51.00	71.00	7.20	1968.00	68511.00	43942.00	53016.00	29513.00	1152.00	52.90
78139	8.15	1881.00	70.00	7.30	2208.00	68375.00	37243.00	53239.00	25497.00	624.00	56.30
78140	8.00	201.00	55.00	7.30	-1.00	64454.00	40007.00	-1.00	-1.00	-1.00	-1.00
78141	8.45	1321.00	62.00	7.40	-1.00	67794.00	42685.00	-1.00	-1.00	-1.00	-1.00
78142	8.10	2290.00	53.00	7.50	-2784.00	68828.00	43381.00	52648.00	28135.00	888.00	55.40
78143	8.00	179.00	52.00	7.40	2989.00	71312.00	45022.00	50854.00	26931.00	864.00	56.40
78144	8.15	1163.00	56.00	7.60	2891.00	64755.00	42035.00	53084.00	28942.00	888.00	58.20
78145	8.15	2185.00	51.00	7.70	3185.00	64818.00	40881.00	52015.00	27906.00	600.00	58.30
78146	8.25	256.00	50.00	7.70	3283.00	65323.00	42072.00	52332.00	28156.00	648.00	59.80
78147	9.30	1128.00	65.00	7.60	-1.00	68478.00	43282.00	-1.00	-1.00	-1.00	-1.00
78148	9.15	1824.00	57.00	7.70	-1.00	65137.00	42154.00	-1.00	-1.00	-1.00	-1.00
78149	9.30	157.00	50.00	7.60	3724.00	62156.00	39838.00	49929.00	26174.00	696.00	-1.00
78150	8.10	1145.00	50.00	7.70	3626.00	63840.00	39816.00	50382.00	26196.00	720.00	60.50
78151	8.35	2024.00	50.00	7.70	3675.00	63796.00	40782.00	50113.00	25855.00	816.00	61.60

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78152	8.10	2829.00	51.00	7.70	3724.00	62526.00	38556.00	48430.00	24784.00	720.00	59.00
78153	8.25	621.00	51.00	7.70	3528.00	62737.00	39917.00	49916.00	25368.00	792.00	64.20
78154	9.20	1404.00	50.00	7.70	-1.00	33369.00	20993.00	-1.00	-1.00	-1.00	-1.00
78155	12.45	2306.00	50.00	7.60	-1.00	60582.00	38540.00	-1.00	-1.00	-1.00	-1.00
78156	8.25	177.00	83.00	7.70	3626.00	61284.00	39635.00	50178.00	26777.00	888.00	59.80
78157	8.15	1018.00	51.00	7.70	3871.00	59719.00	38210.00	48539.00	25035.00	888.00	62.20
78158	8.45	1734.00	52.00	7.70	4116.00	59772.00	39074.00	49056.00	25708.00	600.00	54.30
78159	8.45	2425.00	50.00	7.70	4018.00	60713.00	35547.00	48124.00	23997.00	552.00	57.40
78160	8.15	25.00	50.00	7.60	3773.00	62949.00	36756.00	49093.00	24350.00	600.00	59.80
78161	9.00	568.00	55.00	7.50	-1.00	65488.00	38316.00	-1.00	-1.00	-1.00	-1.00
78162	7.50	1138.00	55.00	7.50	-1.00	62652.00	39070.00	-1.00	-1.00	-1.00	-1.00
78163	8.30	1584.00	56.00	7.50	3920.00	60019.00	38872.00	47023.00	24188.00	624.00	56.70
78164	8.15	2110.00	55.00	7.60	3724.00	58987.00	37634.00	47024.00	24276.00	576.00	58.10
78165	8.15	2710.00	55.00	7.60	3528.00	57116.00	37054.00	45253.00	24412.00	384.00	54.50
78166	8.15	224.00	63.00	7.60	3430.00	59155.00	36311.00	46091.00	23848.00	432.00	57.00
78167	8.20	723.00	85.00	7.70	3381.00	59034.00	36748.00	45580.00	23781.00	408.00	57.00
78168	10.30	1526.00	50.00	7.50	-1.00	57695.00	37509.00	-1.00	-1.00	-1.00	-1.00
78169	10.15	2057.00	50.00	7.70	-1.00	53183.00	30930.00	-1.00	-1.00	-1.00	-1.00
78170	8.30	2495.00	50.00	7.60	2989.00	53537.00	34148.00	46638.00	24210.00	432.00	54.30
78171	8.30	2961.00	52.00	7.60	2891.00	53077.00	34158.00	47032.00	24809.00	408.00	54.60
78172	8.15	413.00	53.00	7.60	2892.00	43435.00	27053.00	46280.00	24167.00	408.00	58.10
78173	8.20	766.00	63.00	7.60	2768.00	49687.00	32606.00	46073.00	24344.00	408.00	55.20
78174	8.40	1290.00	52.00	7.60	2670.00	68788.00	29577.00	44793.00	21753.00	408.00	57.50
78175	9.20	1671.00	53.00	7.60	-1.00	55060.00	36715.00	-1.00	-1.00	-1.00	-1.00
78176	8.50	1827.00	47.00	7.60	-1.00	53713.00	32614.00	-1.00	-1.00	-1.00	-1.00

Table A-1b. START 4-12-78 CORN REACTOR #1 100 L/DAY DET. TIME = 7.75 DAYS

DATE/TIME	GAS METER	FIELD RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78177	8.20	2534.00	51.00	7.50	2597.00	53937.00	32232.00	45448.00	24388.00	288.00	56.20
78178	8.15	2918.00	51.00	7.00	2450.00	47068.00	30549.00	42510.00	22199.00	360.00	57.00
78179	8.05	148.00	50.00	7.50	2450.00	46139.00	29892.00	40200.00	20852.00	240.00	57.00
78180	8.05	569.00	55.00	7.60	2401.00	47940.00	31060.00	39563.00	20543.00	288.00	56.90
78181	8.15	1027.00	63.00	7.50	2254.00	18317.00	10248.00	39593.00	21174.00	360.00	56.80
78182	9.55	1871.00	52.00	-1.00	-1.00	50056.00	38617.00	-1.00	-1.00	-1.00	-1.00
78183	12.50	1938.00	57.00	-1.00	-1.00	44242.00	25019.00	-1.00	-1.00	-1.00	-1.00
78184	8.00	2237.00	50.00	7.50	2156.00	41961.00	24452.00	37082.00	18446.00	312.00	59.90
78185	8.45	2601.00	56.00	7.40	2058.00	49136.00	31901.00	38101.00	19468.00	432.00	55.00
78186	8.15	2901.00	50.00	7.50	2009.00	43044.00	25675.00	37694.00	19306.00	384.00	58.40
78187	8.15	163.00	54.00	7.50	2107.00	39216.00	23446.00	37791.00	19231.00	264.00	56.60
78188	8.10	431.00	57.00	7.50	2009.00	47450.00	30786.00	35421.00	17691.00	312.00	53.10
78189	9.00	652.00	54.00	7.50	-1.00	45148.00	28067.00	-1.00	-1.00	-1.00	-1.00
78190	11.00	903.00	51.00	7.40	-1.00	36815.00	21280.00	-1.00	-1.00	-1.00	-1.00
78191	8.00	1046.00	56.00	7.40	1862.00	41610.00	24617.00	34833.00	17521.00	336.00	54.80
78192	8.00	1127.00	53.00	7.30	1862.00	49841.00	30250.00	34820.00	17528.00	360.00	58.50
78193	8.03	1277.00	52.00	7.30	1813.00	44054.00	26357.00	36576.00	19080.00	456.00	55.20
78194	8.00	1418.00	50.00	7.30	1715.00	19944.00	10622.00	34440.00	17476.00	552.00	58.00
78195	9.00	1558.00	53.00	7.30	1715.00	52315.00	34052.00	33355.00	16473.00	144.00	51.00
78196	11.10	2058.00	49.00	7.10	-1.00	47746.00	31591.00	-1.00	-1.00	-1.00	-1.00

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	APP. T.S.	APP. V.S.	APP. V.A.	CR4 %
78197	12.50	2691.00	50.00	7.00	-1.00	51048.00	33454.00	-1.00	-1.00	-1.00
78198	8.00	110.00	52.00	7.10	1715.00	44730.00	26646.00	34941.00	17963.00	1296.00
78199	8.00	635.00	51.00	7.10	1715.00	40957.00	26512.00	35526.00	18623.00	1176.00
78200	8.30	1204.00	51.00	7.20	1715.00	44750.00	26789.00	34023.00	17791.00	-1.00
78201	8.05	1943.00	52.00	7.30	1862.00	40917.00	27760.00	35713.00	18886.00	1512.00
78202	8.05	2794.00	60.00	7.30	1862.00	47068.00	33106.00	35325.00	18939.00	1032.00
78203	10.30	511.00	54.00	7.40	-1.00	44296.00	33335.00	-1.00	-1.00	-1.00
78204	11.15	1162.00	61.00	7.40	-1.00	40133.00	27648.00	-1.00	-1.00	-1.00
78205	8.10	1749.00	54.00	7.40	2058.00	41199.00	28413.00	33232.00	17573.00	1104.00
78206	8.05	2410.00	52.00	7.30	1960.00	46852.00	23161.00	32958.00	16992.00	1008.00
78207	8.00	43.00	61.00	7.40	2069.00	40777.00	21108.00	31654.00	17359.00	768.00
78208	8.00	669.00	50.00	7.40	2058.00	44970.00	24394.00	32952.00	16611.00	624.00
78209	8.05	1263.00	50.00	7.50	2107.00	47148.00	21357.00	33802.00	16929.00	600.00
78210	8.15	1814.00	55.00	7.40	-1.00	46172.00	24913.00	-1.00	-1.00	-1.00
78211	8.10	2304.00	54.00	7.40	-1.00	44810.00	25427.00	-1.00	-1.00	-1.00
78212	8.15	2355.00	50.00	7.40	2103.00	41134.00	22940.00	31417.00	14979.00	648.00
78213	8.10	452.00	54.00	7.40	2254.00	42336.00	24389.00	33968.00	16614.00	552.00
78214	8.15	922.00	55.00	7.40	2205.00	48363.00	25566.00	33497.00	16030.00	528.00
78215	8.10	1538.00	55.00	7.30	2254.00	48436.00	26255.00	32405.00	15092.00	648.00
78216	8.20	2167.00	57.00	7.30	2152.00	47471.00	25513.00	34090.00	16318.00	672.00
78217	8.35	2719.00	55.00	7.40	-1.00	50078.00	25546.00	-1.00	-1.00	-1.00
78218	11.00	274.00	52.00	7.40	-1.00	45659.00	24005.00	-1.00	-1.00	-1.00
78219	8.00	633.00	54.00	7.50	2401.00	40195.00	21213.00	34389.00	15521.00	504.00
78220	8.05	1111.00	50.00	7.40	2450.00	54595.00	31173.00	35081.00	16178.00	600.00
78221	8.05	1730.00	54.00	7.50	2499.00	53329.00	30832.00	34255.00	23634.00	432.00

DATE/TIME	GAS METER	FIELD H2CL	PH	ALKALINITY	FLUO T.S.	FLUO V.S.	EFF. T.S.	EFF. V.S.	EFF. F.A.	CH4 %	
78222	8.05	2568.00	50.00	7.40	2401.00	53743.00	32739.00	35132.00	16084.00	888.00	55.20
78223	8.05	407.00	48.00	7.50	2322.00	51313.00	32270.00	34917.00	16095.00	528.00	56.50
78224	9.45	1193.00	50.00	7.50	-1.00	49791.00	29901.00	-1.00	-1.00	-1.00	-1.00
78225	11.15	2039.00	53.00	7.50	-1.00	45535.00	26971.00	-1.00	-1.00	-1.00	-1.00
78226	8.05	2652.00	52.00	7.50	2459.00	45974.00	24019.00	35932.00	16550.00	504.00	58.50
78227	8.10	208.00	50.00	7.30	2352.00	44502.00	27569.00	35609.00	16351.00	720.00	58.10
78228	8.15	984.00	56.00	7.50	2499.00	14352.00	13447.00	37578.00	16095.00	432.00	57.10
78229	8.05	1320.00	52.00	7.50	2450.00	17370.00	13596.00	33805.00	15374.00	384.00	58.40
78230	8.20	1631.00	48.00	7.50	2461.00	33844.00	26530.00	31925.00	13673.00	504.00	56.50
78231	7.30	1947.00	56.00	7.40	-1.00	35565.00	34178.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	RFP. T.S.	RFP. V.S.	RFP. V.A.	CH4 %	
78014	10.45	2257.00	50.00	7.20	-1.00	34052.00	24772.00	-1.00	-1.00	-1.00	-1.00
78015	14.30	2712.00	50.00	7.30	-1.00	29800.00	21109.00	-1.00	-1.00	-1.00	-1.00
78016	11.10	34.00	50.00	7.20	1472.00	39608.00	27686.00	26129.00	16335.00	120.00	55.40
78017	8.25	306.00	54.00	7.20	1378.00	35786.00	25785.00	24536.00	14921.00	120.00	54.80
78018	8.20	620.00	55.00	7.20	1425.00	36726.00	26311.00	25034.00	16036.00	120.00	54.10
78019	8.10	960.00	50.00	7.20	1378.00	37665.00	26837.00	26242.00	16917.00	72.00	53.90
78020	8.25	1221.00	52.00	7.20	1282.00	29271.00	19514.00	25209.00	15853.00	72.00	54.80
78021	13.00	1647.00	49.00	7.20	-1.00	32368.00	21000.00	-1.00	-1.00	-1.00	-1.00
78022	11.50	1867.00	49.00	7.00	-1.00	30924.00	21209.00	-1.00	-1.00	-1.00	-1.00
78023	8.15	2177.00	49.00	7.10	1235.00	32716.00	22774.00	24772.00	15495.00	192.00	52.30
78024	8.15	2358.00	53.00	7.20	1282.00	33434.00	23205.00	26330.00	16876.00	96.00	52.30
78025	8.25	2641.00	53.00	7.20	1235.00	32842.00	22789.00	23569.00	14656.00	72.00	52.40
78026	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78027	12.10	192.00	59.00	7.20	1282.00	32193.00	22215.00	24881.00	15630.00	120.00	55.10
78028	10.20	394.00	49.00	7.00	-1.00	33490.00	23269.00	-1.00	-1.00	-1.00	-1.00
78029	12.40	697.00	54.00	7.10	-1.00	33410.00	23285.00	-1.00	-1.00	-1.00	-1.00
78030	8.20	912.00	55.00	7.20	1282.00	29128.00	19407.00	25457.00	15826.00	144.00	52.70
78031	8.10	1171.00	53.00	7.20	1188.00	29648.00	19523.00	24706.00	15328.00	96.00	54.70
78032	8.10	1430.00	54.00	7.20	1235.00	28502.00	18862.00	24723.00	15456.00	96.00	55.30
78033	8.15	1684.00	53.00	7.20	1282.00	30935.00	20578.00	23244.00	13887.00	96.00	54.80
78034	8.15	1941.00	63.00	-1.00	1235.00	29089.00	19682.00	24848.00	15648.00	96.00	54.60
78035	9.35	2212.00	56.00	7.00	-1.00	28454.00	18405.00	-1.00	-1.00	-1.00	-1.00
78036	9.20	2466.00	52.00	7.10	-1.00	28417.00	19175.00	-1.00	-1.00	-1.00	-1.00
78037	8.20	2721.00	76.00	7.20	1330.00	31123.00	21039.00	24174.00	14792.00	120.00	52.40
78038	8.20	2989.00	50.00	7.20	1235.00	34081.00	24435.00	23580.00	14665.00	216.00	51.70

DATE/TIME	GAS METER	PROD RATE	PH	ALKALINITY	PROD T.S.	PROD T.S.	ZPP. T.S.	ZPP. V.S.	ZPP. V.A.	CH4 %	
78039	8.10	159.00	56.00	7.20	1235.00	33754.00	23957.00	24018.00	15000.00	120.00	59.60
78040	8.10	346.00	49.00	7.20	1180.00	32012.00	22173.00	24401.00	15159.00	72.00	53.00
78041	8.15	752.00	49.00	7.10	1188.00	30897.00	21972.00	24570.00	15698.00	72.00	53.00
78042	11.30	1461.00	50.00	7.00	-1.00	37354.00	25715.00	-1.00	-1.00	-1.00	-1.00
78043	1.00	1749.00	54.00	7.20	1235.00	35198.00	23566.00	-1.00	-1.00	-1.00	-1.00
78044	9.10	2044.00	54.00	7.10	1282.00	40964.00	27031.00	24751.00	15555.00	144.00	52.00
78045	8.25	2379.00	61.00	7.20	1282.00	41238.00	23723.00	25479.00	15700.00	96.00	51.40
78046	8.15	2671.00	66.00	7.20	1235.00	43488.00	31017.00	24321.00	14984.00	96.00	53.10
78047	9.15	2976.00	53.00	7.20	1235.00	46177.00	32754.00	26489.00	16874.00	144.00	56.70
78048	9.15	325.00	54.00	7.20	-1.00	38109.00	36566.00	25946.00	16003.00	96.00	54.60
78049	8.40	669.00	51.00	-1.00	-1.00	37840.00	26533.00	-1.00	-1.00	-1.00	-1.00
78050	7.25	987.00	59.00	7.20	-1.00	41740.00	30349.00	-1.00	-1.00	-1.00	-1.00
78051	8.15	1680.00	52.00	7.10	1378.00	40394.00	24862.00	26226.00	16441.00	168.00	52.00
78052	8.15	1974.00	50.00	7.20	1379.00	36513.00	25557.00	24602.00	14997.00	144.00	51.80
78053	8.15	2219.00	71.00	7.20	1330.00	41157.00	29090.00	26050.00	16142.00	144.00	53.20
78054	8.15	2484.00	63.00	7.10	1282.00	35117.00	24319.00	25084.00	15513.00	120.00	52.20
78055	8.20	2725.00	56.00	7.20	1330.00	45202.00	32741.00	23414.00	14227.00	168.00	54.60
78056	9.50	57.00	55.00	7.20	-1.00	45500.00	32126.00	-1.00	-1.00	-1.00	-1.00
78057	9.10	355.00	55.00	7.20	-1.00	43731.00	30992.00	-1.00	-1.00	-1.00	-1.00
78058	4.15	645.00	56.00	7.00	1330.00	42719.00	30346.00	27451.00	17282.00	24.00	51.10
78059	8.00	954.00	47.00	7.20	1330.00	40599.00	26147.00	27771.00	17117.00	384.00	50.70
78060	7.50	1242.00	60.00	7.20	1330.00	39632.00	27717.00	27202.00	16720.00	96.00	53.90
78061	7.35	1522.00	53.00	7.30	1615.00	55434.00	34158.00	28168.00	17077.00	96.00	55.90
78062	7.50	1974.00	53.00	7.20	1472.00	56515.00	30064.00	29399.00	18588.00	384.00	56.80
78063	11.00	2453.00	58.00	7.20	-1.00	60658.00	43475.00	-1.00	-1.00	-1.00	-1.00

Table A-28. COFA STORAGE REACTOR NO. 2 JAN. 74

DATE/TIME	GAS METERS	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78064	11.30	2947.00	59.00	7.10	-1.00	61222.00	43005.00	+1.00	-1.00	-1.00	-1.00
78065	8.20	355.00	55.00	7.20	1425.00	53491.00	33040.00	33401.00	20279.00	192.00	52.40
78066	8.05	822.00	90.00	7.40	1615.00	58816.00	34444.00	32574.00	18810.00	240.00	53.50
78067	11.50	1449.00	78.00	7.40	-1.00	64445.00	43051.00	36126.00	23331.00	+1.00	52.20
78068	8.20	1958.00	78.00	7.20	1805.00	60510.00	42780.00	36003.00	23045.00	168.00	55.70
78069	8.20	2499.00	-1.00	7.30	1805.00	55093.00	32477.00	36261.00	20316.00	168.00	49.00
78070	11.00	125.00	81.00	7.20	-1.00	67391.00	47576.00	-1.00	-1.00	-1.00	-1.00
78071	12.00	755.00	81.00	7.30	-1.00	63924.00	44519.00	-1.00	-1.00	-1.00	-1.00
78072	8.20	1262.00	78.00	7.40	1852.00	59531.00	39787.00	42391.00	27113.00	240.00	52.90
78073	8.15	1939.00	90.00	7.40	1852.00	58789.00	40497.00	39930.00	25323.00	240.00	52.40
78074	8.15	2577.00	80.00	7.40	1852.00	70052.00	46910.00	43460.00	27628.00	216.00	54.80
78075	8.15	280.00	102.00	7.20	1900.00	71916.00	47654.00	44164.00	28036.00	216.00	55.30
78076	8.20	1124.00	85.00	7.30	1948.00	67713.00	35752.00	48132.00	25431.00	264.00	54.80
78077	8.00	1881.00	121.00	7.20	-1.00	62148.00	44300.00	-1.00	-1.00	-1.00	-1.00
78078	9.00	2711.00	90.00	7.30	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78079	8.15	388.00	87.00	7.30	2138.00	64085.00	42819.00	50197.00	31154.00	216.00	53.20
78080	8.05	978.00	80.00	7.30	2208.00	87813.00	50791.00	45611.00	27673.00	192.00	53.20
78081	8.25	1739.00	91.00	7.40	2400.00	77445.00	47228.00	52158.00	31541.00	264.00	55.90
78082	8.10	2529.00	83.00	7.40	2304.00	80704.00	51456.00	52766.00	32047.00	264.00	55.20
78083	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78084	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78085	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78086	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78087	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78088	8.15	2388.00	40.00	7.40	2496.00	82077.00	51780.00	56679.00	32000.00	552.00	46.30

Table A-2a. CORN STOVER REACTOR NO. 2 JAN. 78

DATE/TIME	GAS METER	FIELD RATE	W	AVAILABLE	FUEL T.S.	FUEL V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78089	8.20	70.00	85.00	7.40	2544.00	32907.00	49728.00	61317.00	34873.00	360.00	57.67
78090	8.15	951.00	107.00	7.40	2544.00	41954.00	54569.00	62331.00	35916.00	264.00	54.90
78091	13.32	2349.00	90.00	7.30	-1.00	47121.00	51802.00	-1.00	-1.00	-1.00	-1.00
78092	13.00	417.00	82.00	7.40	-1.00	85715.00	57544.00	-1.00	-1.00	-1.00	-1.00
78093	8.15	1250.00	60.00	7.50	2800.00	46777.00	57663.00	66044.00	38427.00	520.00	54.32
78094	8.10	2139.00	81.00	7.40	2640.00	68325.00	43433.00	66032.00	36562.00	456.00	53.80
78095	3.15	220.00	65.00	7.50	2730.00	66234.00	43742.00	64033.00	37850.00	1968.00	53.60
78096	8.30	1051.00	90.00	7.30	2730.00	64512.00	44114.00	62291.00	36453.00	552.00	52.80
78097	8.20	1835.00	80.00	7.30	2640.00	61754.00	41309.00	52681.00	33711.00	480.00	54.60
78098	8.25	2531.00	50.00	7.30	-1.00	63985.00	36713.00	-1.00	-1.00	-1.00	-1.00
78099	8.15	256.00	61.00	7.40	-1.00	57448.00	18129.00	-1.00	-1.00	-1.00	-1.00
78100	3.30	350.00	92.00	7.50	2928.00	53908.00	37575.00	57701.00	33992.00	384.00	51.40
78101	8.30	1627.00	127.00	7.40	2668.00	61452.00	39912.00	57839.00	32747.00	408.00	53.50

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DATE/TIME	GAS METER	FLED RATE	PH	ALKALINITY	FRED T.S.	FRED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78102	8.30	2409.00	149.00	7.60	2736.00	53831.00	39506.00	56444.00	33844.00	288.00	55.60
78103	8.15	48.00	147.00	7.50	2448.00	65664.00	43857.00	46110.00	25816.00	336.00	55.60
78104	8.25	561.00	138.00	7.30	2160.00	53961.00	39803.00	41412.00	24728.00	312.00	54.10
78105	9.00	1174.00	159.00	7.40	-1.00	56108.00	36108.00	-1.00	-1.00	-1.00	-1.00
78106	10.00	1365.00	145.00	7.10	-1.00	45056.00	31780.00	-1.00	-1.00	-1.00	-1.00
78107	8.15	1873.00	182.00	7.30	1488.00	46796.00	33338.00	35852.00	21655.00	264.00	49.60
78108	8.10	2364.00	159.00	7.20	1488.00	51104.00	36799.00	35388.00	22002.00	264.00	52.10
78109	8.10	2925.00	148.00	7.20	1488.00	53246.00	37526.00	33573.00	20854.00	192.00	51.20
78110	8.25	520.00	150.00	7.00	1344.00	51331.00	35370.00	34154.00	21546.00	192.00	52.30
78111	8.25	1104.00	148.00	7.10	1344.00	59551.00	39802.00	37025.00	23634.00	264.00	53.00
78112	9.45	1806.00	176.00	7.10	-1.00	58504.00	38685.00	-1.00	-1.00	-1.00	-1.00
78113	10.30	2520.00	148.00	7.10	-1.00	57290.00	38441.00	-1.00	-1.00	-1.00	-1.00
78114	8.30	130.00	163.00	7.00	1296.00	58100.00	37680.00	38703.00	23844.00	216.00	56.70
78115	8.05	691.00	152.00	6.90	1227.00	55193.00	36744.00	38557.00	23851.00	168.00	54.40
78116	8.10	1294.00	159.00	7.20	1152.00	54959.00	36095.00	39105.00	24010.00	192.00	46.90
78117	8.30	1886.00	147.00	7.20	1200.00	56424.00	30391.00	41435.00	25724.00	240.00	55.50
78118	8.15	2444.00	149.00	7.10	1200.00	54728.00	29528.00	38125.00	19289.00	168.00	55.60
78119	8.05	-1.00	150.00	7.10	-1.00	48277.00	25680.00	-1.00	-1.00	-1.00	-1.00
78120	9.15	2853.00	154.00	7.10	-1.00	62476.00	43037.00	-1.00	-1.00	-1.00	-1.00
78121	8.25	409.00	147.00	7.20	1104.00	63664.00	41516.00	38380.00	22994.00	264.00	56.00
78122	8.25	917.00	150.00	7.10	960.00	58456.00	36547.00	38167.00	22458.00	216.00	55.30
78123	8.20	1466.00	153.00	7.10	1152.00	63156.00	42382.00	40455.00	24268.00	288.00	52.80
78124	8.25	1800.00	153.00	7.10	1344.00	62230.00	41782.00	42128.00	25529.00	-1.00	51.70
78125	8.15	2260.00	154.00	7.00	1056.00	59199.00	38948.00	40586.00	24223.00	216.00	52.20
78126	9.30	39.00	156.00	7.10	-1.00	60167.00	39032.00	-1.00	-1.00	-1.00	-1.00

Table A-2b. STANT 4-12-78 CORN REACTOR# 2 150 L/DAY DET.FINE=5 DAYS

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	P280 T.S.	P230 V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78127	10.20	720.00	148.00	7.10	-1.00	65258.00	39655.00	+1.00	-1.00	-1.00	-1.00
78128	8.20	1346.00	150.00	7.10	1104.00	62795.00	38353.00	47314.00	28948.00	360.00	49.60
78129	4.10	2041.00	159.00	6.90	1152.00	60095.00	35332.00	48062.00	27548.00	384.00	51.80
78130	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78131	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78132	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78133	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78134	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78135	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78136	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78137	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78138	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78139	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78140	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78141	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78142	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78143	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78144	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78145	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78146	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78147	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78148	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78149	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78150	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78151	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %
78152	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78153	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78154	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78155	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78156	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78157	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78158	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78159	8.45	639.00	30.00	-1.00	-1.00	60713.00	35547.00	-1.00	-1.00	-1.00
78160	8.15	707.00	42.00	-1.00	1754.00	62969.00	36156.00	-1.00	-1.00	912.00
78161	9.00	732.00	48.00	7.20	-1.00	65488.00	38316.00	-1.00	-1.00	-1.00
78162	7.50	1010.00	80.00	7.10	-1.00	62652.00	39070.00	-1.00	-1.00	-1.00
78163	8.30	1484.00	68.00	7.10	1519.00	60219.00	36872.00	44583.00	24315.00	744.00
78164	8.15	2145.00	115.00	7.30	1764.00	58767.00	37634.00	43659.00	23681.00	792.00
78165	8.15	2900.00	66.00	7.30	1911.00	57116.00	37054.00	43326.00	23489.00	744.00
78166	8.15	686.00	66.00	7.40	1911.00	59155.00	36311.00	45190.00	25054.00	62.40
78167	8.20	1461.00	100.00	7.40	2058.00	59034.00	36748.00	44371.00	24781.00	480.00
78168	10.30	2548.00	100.00	7.40	-1.00	57595.00	37504.00	-1.00	-1.00	-1.00
78169	10.15	268.00	103.00	7.40	-1.00	53183.00	30930.00	-1.00	-1.00	-1.00
78170	8.30	1612.00	109.00	7.40	2009.00	53537.00	34148.00	45024.00	24468.00	312.00
78171	8.30	1763.00	103.00	7.40	2009.00	53077.00	34158.00	42528.00	22847.00	288.00
78172	8.15	2579.00	107.00	7.40	2009.00	43435.00	27053.00	42493.00	22609.00	312.00
78173	8.20	156.00	104.00	7.40	2009.00	49687.00	12666.00	40116.00	21200.00	312.00
78174	8.40	857.00	103.00	7.30	1984.00	63788.00	29577.00	40589.00	19862.00	336.00
78175	9.30	1723.00	106.00	7.30	-1.00	55060.00	36715.00	-1.00	-1.00	-1.00
78176	8.50	2655.00	97.00	7.30	-1.00	53713.00	32614.00	-1.00	-1.00	-1.00

Table A-2b. START 4-12-78 COSN REACTOR #2 150 L/DAY 5DAYS

DATE/TIME	GAS METER	FZED RATE	PH	ALKALINITY	FZED T.S.	FZED V.S.	FPP. T.S.	FPP. V.S.	FPP. V.A.	CH4 %
78177										
3.20	574.00	102.00	7.30	2058.00	53937.00	35232.00	40628.00	22619.00	192.00	55.10
78178										
8.15	1536.00	57.00	7.50	2107.00	47060.00	30549.00	40133.00	21810.00	192.00	57.70
78179										
8.05	2190.00	100.00	7.30	2058.00	46139.00	29892.00	39069.00	21155.00	192.00	55.40
78180										
0.05	195.00	104.00	7.40	2058.00	47343.00	31060.00	38824.00	21183.00	192.00	57.10
78181										
8.15	1347.00	101.00	7.40	2107.00	48317.00	32448.00	37694.00	21155.00	166.00	59.90
78182										
9.55	2178.00	101.00	-1.00	-1.00	50056.00	28617.00	-1.00	-1.00	-1.00	-1.00
78183										
12.50	2946.00	102.00	-1.00	-1.00	44242.00	25019.00	-1.00	-1.00	-1.00	-1.00
78184										
8.30	379.00	99.00	7.40	1900.00	41361.00	24452.00	35396.00	18173.00	240.00	58.30
78185										
8.45	1314.00	98.00	7.20	1917.00	49186.00	31901.00	35524.00	18671.00	144.00	53.60
78186										
8.15	1931.00	104.00	7.30	1862.00	43049.00	25675.00	36987.00	19795.00	408.00	56.60
78187										
0.15	2437.00	15.00	7.30	1813.00	39916.00	23446.00	36754.00	19448.00	528.00	55.50
78188										
0.10	2007.00	114.00	7.20	1862.00	47430.00	30796.00	35804.00	19032.00	600.00	52.20
78189										
7.00	319.00	104.00	7.30	-1.00	46148.00	28067.00	-1.00	-1.00	-1.00	-1.00
78190										
11.00	929.00	105.00	7.40	-1.00	36815.00	21280.00	-1.00	-1.00	-1.00	-1.00
78191										
8.00	1241.00	102.00	7.40	1813.00	41610.00	24617.00	34623.00	18537.00	240.00	56.70
78192										
8.00	1006.00	103.00	7.30	1715.00	49141.00	30250.00	34846.00	18038.00	384.00	59.10
78193										
0.05	2160.00	98.00	7.30	1764.00	44054.00	26357.00	35513.00	19053.00	456.00	58.60
78194										
8.00	1607.00	100.00	7.30	1764.00	49944.00	10622.00	30998.00	15743.00	288.00	56.30
78195										
0.00	213.00	100.00	7.30	1813.00	42315.00	34352.00	35927.00	19506.00	552.00	56.50
78196										
11.10	1260.00	102.00	7.20	-1.00	47748.00	31581.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EPP. T.S.	EPP. V.S.	EPP. V.A.	CH4 %	
78197	12.50	2424.00	98.00	7.00	-1.00	51848.00	33454.00	-1.00	-1.00	-1.00	-1.00
78198	8.00	354.00	103.00	7.30	2004.00	44734.00	28646.00	36203.00	19728.00	672.00	61.60
78199	8.00	1420.00	105.00	7.30	2156.00	40457.00	26512.00	35906.00	19898.00	672.00	62.00
78200	8.30	2635.00	102.00	7.40	2205.00	44750.00	28789.00	35829.00	19690.00	624.00	60.30
78201	8.05	887.00	101.00	7.40	2303.00	40317.00	27760.00	34872.00	19122.00	600.00	58.10
78202	8.05	2142.00	100.00	7.40	2303.00	47064.00	33106.00	33391.00	17734.00	552.00	59.70
78203	10.00	349.00	120.00	7.60	-1.00	44296.00	30330.00	-1.00	-1.00	-1.00	-1.00
78204	11.15	1327.00	106.00	7.50	-1.00	40333.00	27648.00	-1.00	-1.00	-1.00	-1.00
78205	8.10	2251.00	105.00	7.40	2352.00	41199.00	28413.00	32534.00	17733.00	840.00	57.20
78206	8.05	318.00	104.00	7.40	2303.00	46852.00	23161.00	32218.00	17037.00	720.00	59.40
78207	8.00	1428.00	102.00	7.40	2303.00	40777.00	21108.00	32568.00	16530.00	624.00	59.50
78208	8.00	60.00	57.00	6.90	784.00	47409.00	27438.00	41048.00	23669.00	120.00	58.40
78209	8.05	414.00	102.00	7.40	2303.00	47148.00	21357.00	32945.00	16094.00	720.00	60.10
78210	8.15	1292.00	107.00	7.40	-1.00	46372.00	24913.00	-1.00	-1.00	-1.00	-1.00
78211	8.10	2169.00	100.00	7.40	-1.00	44630.00	23427.00	-1.00	-1.00	-1.00	-1.00
78212	8.15	47.00	102.00	7.30	2401.00	41134.00	22940.00	33069.00	15739.00	864.00	57.90
78213	8.10	806.00	102.00	7.40	1356.00	42336.00	24389.00	33788.00	16231.00	984.00	56.00
78214	8.15	1598.00	103.00	7.30	2352.00	48063.00	25566.00	34398.00	16374.00	984.00	57.00
78215	8.10	2517.00	101.00	7.20	2401.00	49136.00	26255.00	34381.00	16146.00	1248.00	53.50
78216	8.20	318.00	107.00	7.20	2254.00	47471.00	25513.00	34915.00	16734.00	1296.00	53.90
78217	8.35	993.00	103.00	7.20	-1.00	50078.00	25546.00	-1.00	-1.00	-1.00	-1.00
78218	11.00	1578.00	100.00	7.10	-1.00	45659.00	24005.00	-1.00	-1.00	-1.00	-1.00
78219	8.00	2037.00	96.00	7.20	2205.00	40195.00	21212.00	36051.00	16590.00	1752.00	56.70
78220	8.05	2568.00	104.00	7.00	2254.00	54595.00	31173.00	36840.00	17393.00	2352.00	50.30
78221	8.05	416.00	99.00	6.90	2303.00	53329.00	30832.00	37966.00	17848.00	2328.00	54.40

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DATE/TIME	GAS METER	FEED RATE	pH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78222	8.05	1530.00	101.00	7.00	2205.00	53943.00	32739.00	38426.00	18211.00	2664.00	48.20
78223	8.05	2704.00	100.00	7.10	2352.00	51363.00	32270.00	38645.00	18556.00	2304.00	50.50
78224	9.45	1257.00	99.00	7.40	-1.00	49791.00	29901.00	-1.00	-1.00	-1.00	-1.00
78225	11.15	2418.00	101.00	7.50	-1.00	45035.00	26991.00	-1.00	-1.00	-1.00	-1.00
78226	8.05	208.00	101.00	7.30	2450.00	45378.00	23019.00	37630.00	17868.00	1248.00	57.90
78227	8.10	1032.00	103.00	7.40	2548.00	44542.00	27569.00	38155.00	18149.00	480.00	52.70
78228	8.15	1911.00	26.00	7.40	2665.00	14552.00	13447.00	37385.00	17541.00	1248.00	56.00
78229	8.00	2589.00	61.00	7.40	2499.00	17370.00	13596.00	36416.00	16877.00	1056.00	59.20
78230	8.10	2936.00	50.00	7.40	2401.00	33844.00	26530.00	33140.00	14342.00	1104.00	57.30
78231	7.00	308.00	54.00	7.40	-1.00	35565.00	33178.00	-1.00	-1.00	-1.00	-1.00

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH ₄ %
78014										
10.45	2387.00	56.00	7.20	-1.00	37751.00	27646.00	-1.00	-1.00	-1.00	-1.00
78015										
14.30	2727.00	51.00	7.30	-1.00	33794.00	23509.00	-1.00	-1.00	-1.00	-1.00
78016										
11.10	2968.00	55.00	7.20	1425.30	31282.00	20953.00	22950.00	14065.00	72.00	55.00
78017										
8.25	207.00	50.00	7.20	1070.00	34073.00	23698.00	23005.00	13994.00	96.00	55.60
78018										
8.20	465.00	49.00	7.30	1425.00	35093.00	24721.00	22287.00	13854.00	96.00	55.20
78019										
8.10	723.00	55.00	7.20	1472.00	36513.00	25744.00	23653.00	14789.00	72.00	55.00
78020										
8.25	1011.00	56.00	7.20	1425.00	35532.00	25013.00	22495.00	13749.00	96.00	55.80
78021										
13.00	1330.00	50.00	7.20	-1.00	39510.00	25417.00	-1.00	-1.00	-1.00	-1.00
78022										
11.50	1618.00	48.00	7.10	-1.00	42406.00	30898.00	-1.00	-1.00	-1.00	-1.00
78023										
8.15	1869.00	45.00	7.20	1425.00	40908.00	29206.00	24081.00	14990.00	120.00	53.30
78024										
8.15	2160.00	51.00	7.20	1378.00	41053.00	29278.00	24561.00	15399.00	72.00	53.10
78025										
8.25	2470.00	51.00	7.20	1378.00	42925.00	30521.00	25003.00	15778.00	72.00	52.20
78026										
8.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78027										
12.10	12.00	50.00	7.10	1425.00	46010.00	33499.00	25400.00	15922.00	120.00	55.10
78028										
10.20	203.00	55.00	7.10	-1.00	39935.00	26840.00	-1.00	-1.00	-1.00	-1.00
78029										
12.40	565.00	51.00	7.00	-1.00	51158.00	37324.00	-1.00	-1.00	-1.00	-1.00
78030										
8.20	860.00	56.00	7.20	1425.00	48088.00	34256.00	28382.00	17710.00	144.00	51.10
78031										
8.10	1267.00	47.00	7.20	1425.00	52529.00	37410.00	27426.00	17205.00	96.00	54.30
78032										
8.10	1707.00	54.00	7.30	1472.00	52490.00	37204.00	28096.00	17524.00	96.00	53.50
78033										
8.15	2160.00	55.00	7.20	1425.00	48630.00	34924.00	29053.00	18175.00	96.00	53.40
78034										
8.15	2612.00	54.00	-1.00	1425.00	48066.00	34099.00	30396.00	19498.00	144.00	56.80
78035										
9.35	69.00	52.00	7.20	-1.00	50031.00	35583.00	-1.00	-1.00	-1.00	-1.00
78036										
9.20	481.00	55.00	7.20	-1.00	46544.00	33188.00	-1.00	-1.00	-1.00	-1.00
78037										
8.20	929.00	48.00	7.20	1508.00	52163.00	37365.00	33205.00	21400.00	120.00	52.70
78038										
8.20	1310.00	54.00	7.20	1520.00	45885.00	33105.00	31953.00	20521.00	144.00	50.70

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DATE/TIME	GAS METER	FIELD RATE	PH	ALKALINITY	PROD T.S.	PROD V.S.	EPF. T.S.	EPF. V.S.	EPF. V.A.	CHG %	
78039	8.10	1547.00	51.00	7.30	1168.00	42020.00	30329.00	29300.00	18526.00	144.00	60.60
78040	8.10	1703.00	52.00	7.20	1472.00	49346.00	35600.00	30941.00	19566.00	120.00	56.10
78041	0.15	2152.00	51.00	7.20	1472.00	49955.00	30447.00	29951.00	19270.00	96.00	54.30
78042	11.30	2698.00	52.00	7.20	-1.00	50073.00	36206.00	-1.00	-1.00	-1.00	-1.00
78043	1.00	147.00	54.00	7.25	-1.00	43573.00	30505.00	-1.00	-1.00	-1.00	-1.00
78044	9.10	519.00	55.00	7.10	1520.00	46375.00	31244.00	31666.00	20129.00	144.00	49.40
78045	8.20	989.00	55.00	7.30	1425.00	48369.00	34085.00	31800.00	20131.00	120.00	49.60
78046	0.15	1413.00	55.00	7.20	1425.00	51505.00	35735.00	32994.00	21266.00	120.00	51.00
78047	0.10	1863.00	70.00	7.20	1570.00	53406.00	36205.00	32671.00	21033.00	144.00	51.20
78048	8.15	2375.00	49.00	7.20	1378.00	51183.00	34667.00	32818.00	20531.00	192.00	53.40
78049	8.40	2032.00	59.00	-1.00	-1.00	51814.00	30655.00	-1.00	-1.00	-1.00	-1.00
78050	7.25	347.00	52.00	7.20	-1.00	55373.00	40192.00	-1.00	-1.00	-1.00	-1.00
78051	8.15	840.00	47.00	7.00	1570.00	54985.00	39060.00	32652.00	20376.00	168.00	47.70
78052	0.15	1342.00	65.00	7.20	1378.00	53365.00	30880.00	34421.00	21790.00	192.00	51.30
78053	0.15	1033.00	47.00	7.20	1425.00	53291.00	33290.00	34351.00	21709.00	168.00	51.90
78054	8.15	2260.00	56.00	7.10	1425.00	51277.00	34297.00	34968.00	22131.00	192.00	51.30
78055	8.20	2750.00	40.00	7.20	1378.00	54343.00	42866.00	35236.00	22551.00	216.00	52.70
78056	9.00	231.00	52.00	7.20	-1.00	48133.00	47535.00	-1.00	-1.00	-1.00	-1.00
78057	9.10	726.00	55.00	7.10	-1.00	60940.00	40673.00	-1.00	-1.00	-1.00	-1.00
78058	8.15	1261.00	57.00	7.10	1425.00	44197.00	45062.00	40296.00	25693.00	-1.00	50.20
78059	1.00	1850.00	49.00	7.20	1472.00	62900.00	38550.00	38370.00	21988.00	144.00	49.70
78060	7.30	2397.00	48.00	7.20	1520.00	62050.00	43766.00	41131.00	25778.00	264.00	51.40
78061	7.35	2096.00	54.00	7.30	1265.00	50260.00	35200.00	39131.00	21967.00	72.00	57.00
78062	7.30	426.00	55.00	7.20	1527.00	61538.00	42043.00	40292.00	25201.00	192.00	54.60
78063	11.00	850.00	55.00	7.20	-1.00	64327.00	45457.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH ₄ %
78064										
11.30	-1.00	54.00	7.20	-1.00	62325.00	43144.00	-1.00	-1.00	-1.00	-1.00
78065										
8.20	2094.00	56.00	7.20	1520.00	64379.00	38049.00	40807.00	22554.00	192.00	49.80
78066										
8.05	2650.00	59.00	7.30	1520.00	63788.00	36625.00	42927.00	23045.00	264.00	50.80
78067										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78068										
4.20	814.00	54.00	7.30	1615.00	57642.00	38486.00	43064.00	27000.00	264.00	54.00
78069										
4.20	1409.00	53.00	7.30	1564.00	57198.00	32714.00	43836.00	22936.00	192.00	57.80
78070										
11.00	2053.00	51.00	7.20	-1.00	60794.00	49890.00	-1.00	-1.00	-1.00	-1.00
78071										
12.00	2642.00	49.00	7.30	-1.00	66214.00	45715.00	-1.00	-1.00	-1.00	-1.00
78072										
8.20	75.00	45.00	7.40	1520.00	60575.00	45635.00	45409.00	27555.00	264.00	50.80
78073										
8.15	686.00	49.00	7.30	1662.00	67158.00	48920.00	45610.00	28880.00	432.00	53.00
78074										
8.15	1267.00	49.00	7.30	1615.00	62113.00	46161.00	45577.00	28618.00	216.00	55.20
78075										
4.15	1841.00	63.00	7.20	1662.00	74408.00	50376.00	46705.00	29550.00	240.00	55.30
78076										
8.20	2455.00	60.00	7.20	1615.00	72824.00	41657.00	46647.00	24084.00	144.00	55.40
78077										
8.00	93.00	45.00	7.30	-1.00	70182.00	47981.00	-1.00	-1.00	-1.00	-1.00
78078										
9.00	723.00	47.00	6.80	-1.00	73383.00	50277.00	-1.00	-1.00	-1.00	-1.00
78079										
8.15	1288.00	50.00	7.30	1615.00	71105.00	48680.00	51152.00	32400.00	240.00	50.80
78080										
8.05	1886.00	49.00	7.20	1680.00	73014.00	46357.00	50558.00	31847.00	216.00	52.80
78081										
8.25	2456.00	52.00	7.30	1728.00	65654.00	42478.00	51826.00	32758.00	144.00	52.10
78082										
8.10	402.00	56.00	7.30	1728.00	65026.00	42726.00	50797.00	35063.00	216.00	56.20
78083										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78084										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78085										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78086										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78087										
-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78088										
8.15	2063.00	40.00	7.30	1680.00	65297.00	42327.00	50514.00	30486.00	288.00	44.90

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DATE/TIME	GAS METER	FOLD RATE	PH	ALKALINITY	PE20 T.S.	PE20 V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %
78089										
8.20	2300.00	65.00	7.30	1728.00	67377.00	42741.00	51932.00	31572.00	264.00	59.10
78090										
8.15	2723.00	97.00	7.30	1728.00	66257.00	40137.00	51642.00	31734.00	336.00	56.30
78091										
11.30	243.00	52.00	7.30	-1.00	53462.00	30743.00	-1.00	-1.00	-1.00	-1.00
78092										
13.00	612.00	51.00	7.20	-1.00	49674.00	35229.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CH4 %	
78093	8.15	889.00	40.00	7.30	1728.00	47461.00	33459.00	49198.00	29709.00	288.00	51.30
78094	8.10	1289.00	51.00	7.30	1584.00	48072.00	32420.00	47195.00	26407.00	312.00	52.50
78095	8.15	1659.00	53.00	7.30	1632.00	45348.00	31994.00	46713.00	27742.00	240.00	52.30
78096	8.30	2053.00	64.00	7.20	1584.00	54361.00	37732.00	48506.00	29997.00	264.00	54.20
78097	8.25	2453.00	54.00	7.20	1536.00	48253.00	33188.00	44406.00	26769.00	240.00	52.70
78098	8.25	2819.00	54.00	7.20	-1.00	44138.00	20347.00	-1.00	-1.00	-1.00	-1.00
78099	8.15	206.00	50.00	7.10	-1.00	49497.00	34786.00	-1.00	-1.00	-1.00	-1.00
78100	8.30	585.00	55.00	7.30	1536.00	31200.00	20235.00	44270.00	27446.00	192.00	49.40
78101	8.30	929.00	48.00	7.20	1448.00	59134.00	38041.00	41165.00	24221.00	48.00	53.10
78102	8.30	1247.00	55.00	7.40	1448.00	53475.00	38621.00	41212.00	25048.00	168.00	53.90
78103	8.20	1558.00	54.00	7.30	1392.00	53241.00	17084.00	37613.00	21963.00	216.00	54.10
78104	8.25	1829.00	50.00	7.20	1440.00	47540.00	34557.00	37909.00	23306.00	120.00	53.70
78105	9.00	2140.00	51.00	7.30	-1.00	54556.00	30269.00	-1.00	-1.00	-1.00	-1.00
78106	10.00	2421.00	55.00	7.10	-1.00	48802.00	35510.00	-1.00	-1.00	-1.00	-1.00
78107	8.15	2678.00	54.00	7.20	1448.00	48394.00	34774.00	34931.00	21061.00	168.00	50.00
78108	8.10	2940.00	46.00	7.10	1296.00	49460.00	35138.00	33321.00	20208.00	120.00	52.30
78109	8.10	174.00	60.00	7.10	1248.00	57081.00	34658.00	30707.00	18231.00	144.00	53.60
78110	8.25	403.00	52.00	6.90	1008.00	52730.00	36338.00	30524.00	18713.00	144.00	51.70
78111	8.25	605.00	54.00	7.10	1248.00	53879.00	36746.00	30658.00	18693.00	120.00	56.50
78112	9.45	803.00	60.00	7.20	-1.00	55614.00	37115.00	-1.00	-1.00	-1.00	-1.00
78113	10.30	1245.00	55.00	7.20	-1.00	58341.00	39025.00	-1.00	-1.00	-1.00	-1.00
78114	8.30	1606.00	59.00	7.10	1152.00	53342.00	36680.00	38703.00	23844.00	216.00	56.70
78115	8.05	1982.00	50.00	7.10	1248.00	56177.00	37043.00	35084.00	21286.00	120.00	55.10
78116	8.10	2336.00	67.00	7.20	1248.00	59040.00	38969.00	34084.00	20652.00	192.00	54.40
78117	8.30	2740.00	68.00	7.20	1248.00	54334.00	33107.00	35836.00	21362.00	216.00	56.80

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Table A-3b. CORN STOVES REACTOR #3 JULY 1978

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	2PP. T.S.	2PP. V.S.	2PP. V.A.	CH4 %	
78118	8.15	137.00	56.00	7.20	1296.00	57181.00	33432.00	35680.00	18026.30	240.00	51.20
78119	8.05	441.00	52.00	7.10	-1.00	54560.00	29409.00	-1.00	-1.00	-1.00	-1.00
78120	8.15	644.00	55.00	7.20	-1.00	53312.00	33792.00	-1.00	-1.00	-1.00	-1.00
78121	8.25	1084.00	48.00	7.20	1248.00	59250.00	38567.00	36840.00	21819.00	312.00	56.00
78122	8.25	1404.00	49.00	7.10	1248.00	64534.00	41674.00	37848.00	22165.00	192.00	57.20
78123	8.20	1717.00	48.00	7.20	1248.00	63115.00	42495.00	36723.00	21849.00	216.00	63.20
78124	8.25	2100.00	53.00	7.20	1008.00	64354.00	43663.00	37910.00	22624.00	24.00	54.30
78125	8.15	2308.00	59.00	7.10	1152.00	56574.00	37998.00	34851.00	20322.00	144.00	51.90
78126	9.30	2342.00	54.00	7.20	-1.00	49133.00	32657.00	-1.00	-1.00	-1.00	-1.00
78127	10.20	2868.00	77.00	7.20	-1.00	49227.00	30938.00	-1.00	-1.00	-1.00	-1.00
78128	8.20	142.00	43.00	7.10	1104.00	53290.00	34609.00	35374.00	21062.00	360.00	41.70
78129	8.10	373.00	58.00	7.00	1152.00	50870.00	31448.00	35849.00	20942.00	504.00	51.40
78130	8.10	497.00	47.00	7.00	1152.00	61348.00	34445.00	35877.00	21342.00	504.00	55.00
78131	8.10	652.00	55.00	6.90	1104.00	51541.00	36080.00	36250.00	20990.00	504.00	-1.00
78132	8.15	933.00	53.00	-1.00	1104.00	63351.00	41219.00	37740.00	22589.00	480.00	-1.00
78133	13.25	1515.00	51.00	7.00	-1.00	37647.00	37533.00	-1.00	-1.00	-1.00	-1.00
78134	12.00	1945.00	53.00	7.20	-1.00	53066.00	32738.00	-1.00	-1.00	-1.00	-1.00
78135	8.05	2207.00	55.00	7.00	1152.00	56513.00	32170.00	39181.00	21281.00	192.00	-1.00
78136	8.10	2550.00	50.00	7.10	1200.00	63211.00	40207.00	38821.00	22305.00	288.00	-1.00
78137	8.15	2923.00	82.00	7.10	1200.00	58638.00	36949.00	41221.00	23732.00	384.00	55.00
78138	8.20	305.00	56.00	7.10	1200.00	59813.00	37716.00	41633.00	24131.00	480.00	54.60
78139	8.15	670.00	57.00	7.00	1200.00	59898.00	31920.00	41286.00	21162.00	360.00	54.70
78140	8.00	1043.00	56.00	7.10	-1.00	61759.00	38820.00	-1.00	-1.00	-1.00	-1.00
78141	8.45	1426.00	51.00	7.00	-1.00	64376.00	41614.00	-1.00	-1.00	-1.00	-1.00
78142	8.10	1608.00	51.00	7.20	1248.00	61111.00	39158.00	43645.00	25063.00	336.00	59.50

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	PROD T.S.	PROD V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CR4 %
78143										
8.00	2179.00	59.00	7.70	1323.00	52919.00	31724.00	41258.00	21912.00	312.00	58.40
78144										
8.15	2522.00	51.00	7.30	1323.00	61316.00	40361.00	46026.00	26907.00	624.00	55.20
78145										
8.15	2834.00	59.00	7.20	1470.00	63980.00	40970.00	47598.00	27493.00	264.00	56.50
78146										
8.25	199.00	53.00	7.30	1372.00	60110.00	38342.00	48312.00	27052.00	240.00	55.00
78147										
9.30	616.00	54.00	7.10	-1.00	60954.00	39844.00	-1.00	-1.00	-1.00	-1.00
78148										
9.15	974.00	51.00	7.10	-1.00	62652.00	39511.00	-1.00	-1.00	-1.00	-1.00
78149										
9.30	1335.00	52.00	7.30	1421.00	60189.00	36942.00	47500.00	27279.00	216.00	-1.00
78150										
8.10	1641.00	52.00	7.30	1470.00	63060.00	39057.00	48523.00	27672.00	192.00	57.50
78151										
8.35	2007.00	65.00	7.30	1470.00	61238.00	37968.00	47915.00	26937.00	336.00	60.00
78152										
8.10	2194.00	53.00	7.20	1470.00	59038.00	36255.00	44756.00	26918.00	192.00	63.50
78153										
8.25	2355.00	55.00	7.40	1421.00	53292.00	31911.00	45646.00	25196.00	168.00	64.60
78154										
9.20	2509.00	60.00	7.20	-1.00	22490.00	11856.00	-1.00	-1.00	-1.00	-1.00
78155										
12.45	2794.00	50.00	7.10	-1.00	64461.00	40949.00	-1.00	-1.00	-1.00	-1.00
78156										
8.25	31.00	52.00	7.20	1519.00	61623.00	39483.00	50864.00	29719.00	192.00	60.00
78157										
8.15	312.00	54.00	7.10	1519.00	60951.00	38424.00	49781.00	28538.00	192.00	65.10
78158										
8.45	622.00	64.00	7.30	1470.00	68005.00	41970.00	50213.00	28913.00	192.00	56.00
78159										
8.45	1000.00	52.00	7.20	1617.00	65760.00	37524.00	49372.00	26006.00	192.00	57.70
78160										
8.15	1335.00	51.00	7.20	1617.00	64597.00	37421.00	53168.00	27411.00	168.00	60.30
78161										
9.00	1054.00	48.00	7.20	-1.00	69980.00	41499.00	-1.00	-1.00	-1.00	-1.00
78162										
7.50	1904.00	46.00	7.20	-1.00	70065.00	44231.00	-1.00	-1.00	-1.00	-1.00
78163										
8.30	2174.00	51.00	7.20	1519.00	71252.00	45588.00	46757.00	25719.00	168.00	58.40
78164										
8.15	2413.00	51.00	7.20	1519.00	68005.00	40244.00	46685.00	25934.00	144.00	59.60
78165										
8.15	2699.00	49.00	7.20	1568.00	62152.00	38998.00	47460.00	26514.00	168.00	59.60
78166										
8.15	2995.00	49.00	7.30	1568.00	65702.00	41975.00	41183.00	20791.00	168.00	59.50
78167										
8.20	277.00	51.00	7.30	1519.00	64533.00	42495.00	42142.00	22952.00	144.00	57.70

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DATE/TIME	GAS METEN	FED RATE	PH	ALKALINITY	FED T.S.	FED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CHG %
78168	10.30	606.00	54.00	7.20	-1.00	50240.00	35651.00	-1.00	-1.00	-1.00
78169	10.15	932.00	50.00	7.30	-1.00	57040.00	35494.00	-1.00	-1.00	-1.00
78170	8.30	1255.00	61.00	7.30	1372.00	59073.00	38528.00	45595.00	25535.00	120.00
78171	8.30	1645.00	56.00	7.20	1421.00	59446.00	38186.00	44787.00	24956.00	144.00
78172	8.15	2006.00	58.00	7.20	1227.00	51161.00	29733.00	45376.00	25229.00	144.00
78173	8.20	2354.00	50.00	7.20	1421.00	55422.00	36489.00	46705.00	27015.00	144.00
78174	8.40	2721.00	53.00	7.20	1421.00	53992.00	30751.00	46085.00	23948.00	312.00
78175	9.30	106.00	52.00	7.10	-1.00	59198.00	40015.00	-1.00	-1.00	-1.00
78176	8.50	478.00	53.00	7.20	-1.00	49883.00	32614.00	-1.00	-1.00	-1.00
78177	8.20	856.00	54.00	7.10	1123.00	56183.00	38014.00	47928.00	28427.00	192.00
78178	8.15	1254.00	74.00	7.20	1372.00	55517.00	35956.00	46480.00	26160.00	216.00
78179	8.35	1669.00	54.00	7.00	1274.00	51457.00	34274.00	44015.00	24937.00	168.00
78180	8.05	2024.00	168.00	7.00	1274.00	35023.00	23796.00	43039.00	26438.00	192.00
78181	8.15	2388.00	50.00	7.10	1127.00	60326.00	41644.00	39714.00	25355.00	240.00
78182	9.55	2661.00	45.00	-1.00	-1.00	53106.00	35228.00	-1.00	-1.00	-1.00
78183	12.50	2852.00	50.00	-1.00	-1.00	50099.00	32847.00	-1.00	-1.00	-1.00
78184	8.30	2922.00	51.00	7.00	1029.00	50241.00	33737.00	34637.00	19226.00	192.00
78185	8.45	237.00	52.00	6.90	1029.00	49473.00	32243.00	35947.00	20335.00	168.00
78186	8.15	495.00	51.00	7.10	980.00	49433.00	33661.00	37078.00	21372.00	168.00
78187	8.15	719.00	54.00	7.00	1029.00	48297.00	32617.00	47028.00	32617.00	144.00
78188	8.15	1013.00	50.00	7.10	1029.00	28134.00	19588.00	34760.00	19875.00	168.00
78189	9.00	1217.00	52.00	7.00	-1.00	55602.00	36645.00	-1.00	-1.00	-1.00
78190	11.00	1364.00	50.00	7.00	-1.00	49785.00	32138.00	-1.00	-1.00	-1.00
78191	8.00	1657.00	57.00	7.10	980.00	50574.00	33275.00	36997.00	21686.00	144.00
78192	8.00	1901.00	50.00	7.00	1029.00	29822.00	20044.00	38022.00	22209.00	216.00

DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EPP. T.S.	EPP. V.S.	EPP. V.A.	CH4 %	
78193	8.05	2064.00	59.00	6.90	1029.00	60306.00	39213.00	37721.00	22337.00	168.00	58.60
78194	8.00	2142.00	50.00	6.90	980.00	54425.00	37585.00	42004.00	25559.00	168.00	59.00
78195	8.00	2484.00	57.00	7.00	931.00	57657.00	38137.00	38317.00	22649.00	672.00	55.20
78196	11.10	2871.00	62.00	6.80	+1.00	56141.00	36269.00	-1.00	-1.00	-1.00	-1.00
78197	12.50	190.00	70.00	6.90	-1.00	49592.00	31006.00	-1.00	-1.00	-1.00	-1.00
78198	8.00	428.00	56.00	7.00	833.00	51840.00	34034.00	40260.00	23715.00	192.00	62.00
78199	8.00	759.00	50.00	6.90	931.00	51813.00	32818.00	40675.00	23858.00	96.00	57.20
78200	8.30	1101.00	56.00	6.90	882.00	46648.00	29800.00	41053.00	24169.00	168.00	52.50
78201	8.05	1373.00	60.00	6.90	980.00	50317.00	32964.00	40664.00	24033.00	120.00	51.10
78202	8.05	1629.00	60.00	6.90	882.00	48338.00	31168.00	41897.00	25221.00	168.00	57.70
78203	10.30	1929.00	55.00	7.00	+1.00	54563.00	33805.00	-1.00	-1.00	-1.00	-1.00
78204	11.15	2213.00	57.00	6.90	-1.00	46432.00	29120.00	-1.00	-1.00	-1.00	-1.00
78205	8.10	2437.00	49.00	6.90	882.00	47264.00	29916.00	40727.00	23936.00	144.00	60.90
78206	8.05	2659.00	50.00	6.90	833.00	38245.00	24325.00	41895.00	24540.00	144.00	57.80
78207	8.00	2863.00	56.00	6.90	833.00	53157.00	30115.00	42793.00	25380.00	144.00	57.60
78208	8.00	60.00	57.00	6.90	784.00	47409.00	27438.00	41044.00	23669.00	120.00	58.40
78209	8.05	267.00	53.00	6.90	784.00	49443.00	25803.00	41656.00	22710.00	192.00	57.20
78210	8.15	489.00	56.00	6.80	-1.00	52854.00	28560.00	-1.00	-1.00	-1.00	-1.00
78211	8.10	735.00	51.00	6.90	-1.00	49695.00	26255.00	-1.00	-1.00	-1.00	-1.00
78212	8.15	963.00	50.00	6.80	735.00	32788.00	17496.00	39191.00	21426.00	120.00	56.70
78213	8.10	1178.00	55.00	6.80	686.00	49653.00	29529.00	41787.00	23643.00	168.00	57.00
78214	8.15	1408.00	54.00	6.80	735.00	48459.00	28589.00	40591.00	22412.00	144.00	57.70
78215	8.10	1608.00	58.00	6.80	735.00	49329.00	28188.00	41833.00	23133.00	192.00	57.10
78216	8.20	1890.00	53.00	6.70	784.00	49378.00	28463.00	42652.00	24582.00	168.00	56.70
78217	8.35	2119.00	57.00	6.80	-1.00	55444.00	31846.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS WATER	FEED RATE	PH	ALKALINITY	FEED T.S.	PERO T.S.	ZPP, T.S.	EFF. T.S.	ZPP, V.A.	CH4 %
78210										
11.00	2371.00	57.00	6.80	-1.00	53749.00	30549.00	-1.00	-1.00	-1.00	-1.00
78219										
8.00	2525.00	47.00	6.90	733.00	51032.00	29414.00	41842.00	23031.00	216.00	58.50
78220										
8.35	2624.00	59.00	6.70	686.00	51394.00	30642.00	41499.00	22180.00	336.00	49.10
78221										
8.05	2808.00	52.00	6.70	686.00	48031.00	23129.00	42578.00	15892.00	408.00	53.60
78222										
8.35	66.00	54.00	6.80	637.00	47961.00	27944.00	40689.00	22229.00	144.00	50.30
78223										
8.05	190.00	51.00	6.80	686.00	52098.00	32322.00	42928.00	24070.00	360.00	54.50
78224										
9.45	414.00	50.00	6.90	-1.00	50242.00	31646.00	-1.00	-1.00	-1.00	-1.00
78225										
11.15	685.00	53.00	6.90	-1.00	45454.00	27267.00	-1.00	-1.00	-1.00	-1.00
78226										
8.35	891.00	51.00	6.80	686.00	46066.00	28541.00	41976.00	23068.00	216.00	57.80
78227										
8.10	1057.00	48.00	6.70	637.00	40946.00	24528.00	41776.00	22990.00	312.00	54.60
78228										
8.15	1145.00	56.00	6.70	686.00	43876.00	26232.00	41894.00	22976.00	432.00	51.00
78229										
8.35	1269.00	54.00	6.70	583.00	40123.00	24132.00	42017.00	23349.00	384.00	56.30
78230										
8.10	1441.00	58.00	6.80	583.00	39434.00	28588.00	40684.00	19961.00	312.00	54.50
78231										
7.00	1630.00	54.00	6.80	-1.00	7694.00	3865.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GRS RETEN	FEED RATE	PH	ALKALINITY	PROD T.S.	PROD V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CBH %	
78014	13.45	201.00	51.00	7.20	-1.00	37751.00	27646.00	-1.00	-1.00	-1.00	-1.00
78015	14.30	577.00	54.00	7.30	-1.00	33794.00	23509.00	-1.00	-1.00	-1.00	-1.00
78016	11.10	834.00	51.00	7.20	1425.00	31282.00	20953.00	25983.00	15759.00	120.00	57.10
78017	8.25	1094.00	47.00	7.20	1425.00	34073.00	23698.00	24336.00	14703.00	72.00	55.70
78018	8.20	1380.00	50.00	7.20	1425.00	35293.00	24721.00	25692.00	16282.00	192.00	55.20
78019	8.10	1671.00	50.00	7.20	1425.00	36513.00	25744.00	25674.00	16146.00	96.00	54.80
78020	8.25	1968.00	60.00	7.20	1330.00	35532.00	25013.00	26251.00	16524.00	72.00	55.20
78021	13.00	2334.00	52.00	7.25	-1.00	39510.00	25417.00	-1.00	-1.00	-1.00	-1.00
78022	11.50	2603.00	45.00	7.20	-1.00	42408.00	30898.00	-1.00	-1.00	-1.00	-1.00
78023	8.15	2673.00	54.00	7.20	1379.00	40908.00	29206.00	27284.00	17288.00	72.00	52.70
78024	8.15	189.00	52.00	7.20	1425.00	41053.00	29278.00	26497.00	16551.00	96.00	53.10
78025	8.25	525.00	52.00	7.20	1425.00	42925.00	30521.00	25281.00	15570.00	144.00	49.80
78026	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00	-1.00
78027	12.10	1121.00	47.00	7.10	1425.00	46610.00	31499.00	26846.00	17292.00	120.00	54.70
78028	10.20	1327.00	53.00	7.10	-1.00	39935.00	26840.00	-1.00	-1.00	-1.00	-1.00
78029	12.40	1699.00	58.00	6.90	-1.00	51158.00	37324.00	-1.00	-1.00	-1.00	-1.00
78030	8.20	2011.00	55.00	7.20	1475.00	48088.00	34256.00	30005.00	18851.00	192.00	51.50
78031	8.10	2445.00	50.00	7.20	1425.00	52529.00	37410.00	28699.00	17952.00	96.00	54.70
78032	8.10	2882.00	50.00	7.20	1472.00	52490.00	37204.00	29675.00	18730.00	120.00	51.80
78033	8.15	312.00	52.00	7.20	1425.00	48830.00	34924.00	28877.00	18004.00	120.00	52.50
78034	8.15	722.00	49.00	-1.00	1472.00	48066.00	34099.00	30403.00	19168.00	120.00	57.20
78035	9.35	1143.00	57.00	7.20	-1.00	50011.00	33583.00	-1.00	-1.00	-1.00	-1.00
78036	9.20	1594.00	61.00	7.10	-1.00	46544.00	33188.00	-1.00	-1.00	-1.00	-1.00
78037	8.20	1953.00	63.00	7.20	1568.00	52163.00	37365.00	31517.00	19679.00	96.00	53.90
78038	8.20	2916.00	53.00	7.20	1520.00	45085.00	33105.00	31953.00	20521.00	144.00	50.70

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CR4 %	
78039	8.10	2700.00	51.00	7.30	1568.00	42026.00	30329.00	32661.00	20881.00	144.00	60.10
78040	8.10	2953.00	52.00	7.20	1568.00	49346.00	35600.00	33098.00	21065.00	96.00	55.60
78041	8.15	363.00	50.00	7.20	1520.00	49355.00	36447.00	33180.00	21222.00	140.00	52.00
78042	11.30	936.00	55.00	7.10	-1.00	50673.00	36206.00	-1.00	-1.00	-1.00	-1.00
78043	1.00	1343.00	62.00	7.20	-1.00	43373.00	39535.00	-1.00	-1.00	-1.00	-1.00
78044	9.10	1708.00	54.00	7.10	1472.00	46735.00	31244.00	32584.00	20577.00	144.00	49.60
78045	8.25	2198.00	62.00	7.00	1520.00	48369.00	34085.00	31426.00	19659.00	144.00	49.70
78046	8.15	2678.00	68.00	7.30	1573.00	51305.00	35735.00	30322.00	18711.00	144.00	48.30
78047	8.15	136.00	57.00	7.20	1578.00	53466.00	33285.00	33359.00	21403.00	648.00	48.70
78048	8.15	690.00	46.00	7.20	1378.00	51183.00	44667.00	31447.00	18934.00	168.00	48.20
78049	8.40	1476.00	59.00	-1.00	-1.00	51014.00	36655.00	-1.00	-1.00	-1.00	-1.00
78050	7.25	1677.00	57.00	7.10	-1.00	55373.00	40192.00	-1.00	-1.00	-1.00	-1.00
78051	8.15	2256.00	49.00	7.00	1262.00	54985.00	33060.00	33843.00	21220.00	408.00	46.60
78052	8.15	2827.00	65.00	7.20	1330.00	53360.00	30883.00	32541.00	19878.00	360.00	50.10
78053	8.15	338.00	54.00	7.20	1330.00	53291.00	30290.00	32642.00	20136.00	408.00	50.40
78054	8.15	793.00	54.00	7.10	1250.00	51277.00	34297.00	33473.00	20721.00	408.00	50.40
78055	8.20	1138.00	51.00	7.20	1130.00	59343.00	42866.00	33761.00	21158.00	552.00	45.60
78056	9.50	1305.00	71.00	6.90	-1.00	68133.00	47535.00	-1.00	-1.00	-1.00	-1.00
78057	7.10	1764.00	53.00	6.90	-1.00	60948.00	40673.00	-1.00	-1.00	-1.00	-1.00
78058	8.15	2029.00	65.00	6.80	1378.00	64197.00	45662.00	37530.00	31362.00	3504.00	45.90
78059	8.00	2560.00	54.00	6.90	1330.00	62400.00	38550.00	40732.00	23793.00	1344.00	50.30
78060	7.50	73.00	58.00	7.00	1378.00	62450.00	43760.00	36702.00	22160.00	744.00	54.80
78061	7.35	772.00	55.00	7.30	1472.00	58200.00	35204.00	38255.00	21315.00	120.00	53.40
78062	7.50	1287.00	59.00	7.20	1520.00	61530.00	42043.00	37751.00	22823.00	528.00	55.10
78063	11.00	1666.00	55.00	7.20	-1.00	64027.00	45457.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	PLD T.S.	FEED V.S.	EPP. T.S.	EPP. V.S.	EPP. V.A.	CH4 %	
78064	11.30	1985.00	55.00	7.10	-1.00	62325.00	43144.00	-1.00	-1.00	-1.00	-1.00
78065	8.20	2272.00	49.00	7.10	1425.00	64379.00	38049.00	42048.00	23827.00	1152.00	49.80
78066	8.05	2598.00	66.00	7.00	1425.00	63780.00	36625.00	40912.00	21280.00	1440.00	41.80
78067	11.50	3026.00	48.00	6.80	-1.00	65647.00	41287.00	43708.00	26545.00	+1.00	48.20
78068	8.20	456.00	54.00	7.00	1425.00	57042.00	38486.00	42085.00	26150.00	1488.00	54.00
78069	8.20	1040.00	55.00	7.10	1520.00	57188.00	32714.00	43028.00	22046.00	1296.00	56.70
78070	11.00	1554.00	67.00	7.40	-1.00	68794.00	49890.00	-1.00	-1.00	-1.00	-1.00
78071	12.00	2117.00	49.00	7.40	-1.00	60214.00	45715.00	-1.00	-1.00	-1.00	-1.00
78072	8.20	2857.00	51.00	7.40	1520.00	68575.00	43635.00	44878.00	27634.00	720.00	50.40
78073	8.15	191.00	54.00	7.30	1615.00	67158.00	48920.00	44841.00	28139.00	720.00	56.10
78074	8.15	491.00	55.00	7.20	1520.00	62113.00	46161.00	46833.00	29951.00	912.00	56.70
78075	8.15	896.00	50.00	7.00	1520.00	74408.00	50376.00	44988.00	28143.00	984.00	52.10
78076	8.20	1611.00	75.00	7.00	1478.00	72824.00	41657.00	48470.00	25421.00	816.00	53.40
78077	8.00	2595.00	53.00	7.00	-1.00	70182.00	47981.00	-1.00	-1.00	-1.00	-1.00

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DATE/TIME	GAS METER	FEED RATE	PH	ALKALINITY	FEED T.S.	FEED V.S.	EFF. T.S.	EFF. V.S.	EFF. V.A.	CHG %
78099	8.15	206.00	56.00	7.10	-1.00	49497.00	34786.00	-1.00	-1.00	-1.00
78100	8.30	1796.00	40.00	7.30	1584.00	31200.00	20235.00	44896.00	27060.00	1200.00
78101	8.30	1899.00	117.00	7.30	1584.00	59134.00	38041.00	42882.00	24315.00	600.00
78102	8.30	2103.00	168.00	7.30	1536.00	53475.00	38621.00	43297.00	26056.00	408.00
78103	8.20	2519.00	198.00	7.20	1440.00	53241.00	37084.00	40790.00	24451.00	696.00
78104	8.25	37.00	193.00	7.10	1248.00	47540.00	34557.00	36847.00	22903.00	504.00
78105	9.00	552.00	220.00	7.10	-1.00	54556.00	33269.00	-1.00	-1.00	-1.00
78106	10.00	1086.00	207.00	6.90	-1.00	48802.00	35510.00	-1.00	-1.00	-1.00
78107	8.15	1572.00	202.00	7.00	1206.00	48394.00	34774.00	35513.00	22593.00	408.00
78108	8.10	2149.00	203.00	7.00	1056.00	49860.00	35138.00	32385.00	20821.00	336.00
78109	8.10	2627.00	198.00	7.00	1056.00	57381.00	38658.00	32899.00	21337.00	408.00
78110	8.25	173.00	196.00	6.80	1290.00	52730.00	36338.00	33069.00	21197.00	432.00
78111	8.25	758.00	199.00	6.90	1056.00	53879.00	36746.00	34877.00	22349.00	408.00
78112	9.45	1473.00	194.00	7.00	-1.00	55618.00	37115.00	-1.00	-1.00	-1.00
78113	10.30	2220.00	200.00	7.00	-1.00	58341.00	39025.00	-1.00	-1.00	-1.00
78114	8.30	2655.00	124.00	6.90	1152.00	53942.00	35094.00	38833.00	24201.00	384.00
78115	8.05	522.00	157.00	7.00	1200.00	56177.00	37043.00	40107.00	25037.00	240.00
78116	8.10	1212.00	203.00	7.10	1152.00	59040.00	38969.00	40105.00	25070.00	408.00
78117	8.30	1339.00	202.00	7.10	1200.00	54334.00	33107.00	40162.00	24804.00	552.00
78118	8.15	2679.00	192.00	7.00	1200.00	57161.00	32432.00	42603.00	21967.00	408.00
78119	8.45	454.00	204.00	7.10	-1.00	54560.00	29409.00	-1.00	-1.00	-1.00
78120	8.25	1495.00	202.00	7.10	-1.00	53312.00	33792.00	-1.00	-1.00	-1.00

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APPENDIX 8

ANALYTICAL TECHNIQUES

1. Wet Sieve Analysis

The wet sieve analysis was used to separate a sample of slurry into various size fractions.

Apparatus:

Standard sieves - sizes:

- #10 - 1.981 mm
- #20 - 0.840 mm
- #50 - 0.297 mm
- #100 - 0.149 mm
- #200 - 0.074 mm

Procedure:

1. Dry and tare sieves to within ± 5 grams.
2. Collect and measure 2 liters of slurry.
3. Allow slurry to settle for at least 10 minutes.
4. Pass supernatant through #10 and #20 mesh sieves. Rotate sieves to encourage passage of solids. Transfer settled solids to screen and allow to drain.
5. Set liquid aside and allow solids to settle for 15 to 20 minutes.
6. Wash solids on first 2 screens with small quantities of water. Wash until clean water passes screen (less than 4 liters). Collect wash water.
7. Pass wash water through the remaining screens. Collect wash water and use to wash accumulated solids on the remaining sieves.
8. Pour supernatant from step 5 through the next 2 finer screens. Collect liquid. Transfer settled solids to screens and allow to drain.
9. Using wash water from step 7, wash accumulated solids.
10. Pour remaining liquid to other screens and repeat steps 8 and 9.
11. After above steps have been performed for all sieves, collect and measure volume of passed liquid and wash water.
12. Allow solids on screens to dry in 103°C oven and weigh when dry.

2. Fiber Analysis

This test measures the hemicellulose and cellulose as holocellulose. The fats, oils, waxes, etc., are extracted with an organic solvent. Acid hydrolysis removes the holocellulose and the remaining solids are ignited at 600°C to measure the inorganic solids.

Apparatus:

Soxhlet extraction apparatus

250 ml soxhlet extraction flask

100 ml soxhlet extraction tube

Cool water condensers

25 x 80 mm cellulose crucibles, hot plate

1 liter side arm erlenmeyer flasks

25 ml Gooch crucibles w/asbestos pads

32 mm filtration holders for Gooch crucibles

125 ml erlenmeyer flasks

Reagents:

Alcohol-benzene solution (1 volume ethyl alcohol to 2 volumes of c.p. benzene), sulfuric acid (72 percent)

Procedure:

1. Grind test samples to a size of less than 1 mm by passing it through a Wiley Mill.
2. Determine percent fixed solids of sample material. Dry and weigh a crucible without lid on it. Place dry sample in crucible and reweigh. The difference gives the weight of the sample. Combust the sample at 600°C for 2-3 hours with the lid on. Remove the lid from the crucible and let stand at 600°C for 4-5 hours for complete combustion. Cool the crucible in a desiccator and weigh it without the lid. Subtract the weight of the crucible to obtain the fixed solids remaining. Divide this remaining weight by the initial weight of the sample to obtain the percentage fixed solids in the sample.
3. Place 120 ml of alcohol-benzene solution in extraction flask. Connect flask, tube with cellulose crucibles, and condensor and

place in water bath. Extract empty crucibles for 4 hours to remove any extractables in the crucible.

4. Dry and weigh empty crucibles. Fill crucibles with sample and dry and weigh. Difference gives the weight of the sample. Follow extraction procedure as given for empty crucibles.
5. Dry and weigh crucibles and extracted sample. Weight loss is the weight of the extractable material.
6. Dry and weigh 125 ml flasks. Add approximately 1 g of extracted sample of flask; dry and weigh again. Add 15 ml of cold sulfuric acid to flask while stirring. Place flask in cool water bath (20°C) for 2 hours and stir frequently. Wash material from flask into 1 liter beaker with distilled water. Add 500 ml of distilled water to beaker and boil for 4 hours keeping volume constant.
7. Dry and weigh Gooch crucibles with asbestos pad. Filter beaker contents through Gooch crucibles. Dry and weigh crucibles. Material in crucible is lignin and fixed solids.
8. Subtract percent of fixed solids from percent of lignin and fixed solids.

As in step 7, the material not only contains lignin and fixed solids, but protein present in the residue precipitates and is measured as part of the residue which is designated as lignin. Total nitrogen of the residue is determined to correct for protein which is 6.25 times total nitrogen. Therefore, net lignin = (weight in step 7) - fixed solids - protein.

3. Filter Leaf Test Procedure

The test procedure discussed below has, for the most part, been followed for the entire testing period. Slight variations have been made to accommodate various peculiarities of the sludge in question and to ease the operation of the test. Deviations from this method of testing have been mentioned previously. These deviations are minor and were designed in such a way that their effect on the continuity of the data is minimal.

The tests were run on actual fresh digested material. In no instance was the digester effluent older than 5 hours. A 5 gallon sample was obtained from the digester effluent tanks while the effluent was subjected to compressed air agitation. For each test run a known volume of

sample (1 to 3 liter depending on the substrate) was obtained under vigorous mechanical mixing from the larger volume.

Much of testing procedure involves determining the percentage of solids by weight of a given material. Since evaporative loss of weight techniques were used, the handling and weighing of crucibles was necessary. Therefore, a consistent and theoretically sound method of crucible weighing and handling had to be determined. First this technique is enumerated and then the actual filter leaf test will be described.

Crucible Handling Technique:

1. Wash crucible with soap and water using spatulas and brushes to remove foreign objects. Rinse with distilled water.
2. Crucibles are placed in muffle furnace (600°C) to oxidize any organic matter remaining.
3. Crucibles are moved to drying oven (103°C) to cool.
4. Cooled crucibles are dry brushed out to remove ash formed in step 2.
5. Crucibles are stored in drying oven in order to avoid the additional weight of adsorbed water.
6. When a crucible is to be weighed, it is placed in a desiccator for about 45 minutes. (Crucibles must be cool to avoid convection currents during weighing. The crucibles are stored in desiccator to keep water from adsorbing on them.)
7. Crucibles are weighed to 10^{-3} grams.
8. Crucibles that are used in a weight added mode are stored in atmospheric conditions.
9. Crucibles used where the total weight of crucible and added material is determined are stored in a drying oven until use.

Crucibles are not touched by hand before or after weighing.* An exception to this is where the tare weight is known and a known weight is

*Since oils and moisture transferred from hand to crucible add an unaccounted foreign weight to crucibles.

added to the crucible. Any oils transferred to the crucible after taring will be volatilized off in the oven before the next weighing.

Filter Leaf Test Procedure:

A step by step procedure is listed below. This procedure is for one run. However, step 4 and step 5 are needed only once every 4 runs.

1. Wash test equipment
2. Rinse test equipment (except hoses and filter support) with distilled water.
3. Select 1 Gooch crucible and 4 tared crucibles.
4. Agitate anaerobic effluent tank with compressed air mixer.
5. Immerse clean 5 gallon bucket in mixed effluent tank to obtain representative sample.
6. Set up mechanical mixer in 5 gallon bucket and turn mixer on.
7. Take representative 150 ml sample from well mixed 5 gallon bucket.
8. While stirring 150 ml sample, pour 60 g of liquor into crucible #1. (Determine weight to 0.25 g and record.) Place in drying oven.
9. Take the remainder of the 150 ml sample and record its pH to 10^{-1} accuracy.
10. Withdraw 3 liters from the well mixed 5 gallon bucket and transfer to the clean 4 liter beaker.
11. Turn off rapid mix. (Continuous mixing causes CO_2 release, thus raising pH.)
12. Turn on vacuum pump to preset pressure. (Usually 16 inches of mercury.)
13. Turn off vacuum actuate and vacuum release valve allowing pressure to build up in 5 gallon vacuum reservoir.
14. Measure out coagulant or polymer dose (if any).
15. Place mechanical mixer in 4 liter beaker.
16. Turn on mixer (go to next step without delay).
17. At time = 0 minutes turn on stopwatch, dump coagulant or polymer into 4 liter beaker.

18. At time = 0.5 minutes, turn off rapid mix.
19. Transfer 4 liter beaker to magnetic stirrer.
20. Record pH.
21. At time = 6.5 minutes, turn on magnetic stirrer to low setting (#2).
22. At time = 7.75 minutes, place filter leaf holder in a 4 liter beaker.
23. At time = 7.0 minutes, open vacuum actuate valve (form time begins).
24. Record form pressure 15 seconds into the form time.
25. At the end of the form time remove the filter leaf from the 4 liter beaker and place it upright in the holder (dry time begins).
26. Turn off magnetic stirrer.
27. 15 seconds into the dry time record the drying pressure.
28. At the end of dry time turn off the vacuum pump and open the vacuum reservoir release valve.
29. Remove cake from test leaf with spatula.
30. Record thickness of cake.
31. Place cake in crucible number 2, weigh cake and crucible, place in drying oven.
32. Empty filtrate in vacuum flask into 1 liter graduated cylinder.
33. Record volume.
34. Pour known weight of filtrate into crucible number 3 and determine weight to 0.25 g. Place in drying oven.
35. Pour 70 g of filtrate into centrifuge tube using balance (determine weight to 0.25 g).
36. Centrifuge sample at 4500 rpm for 40 minutes.
37. Turn on second vacuum pump.
38. Wet glass filter in Gooch crucible with distilled water.
39. Subject Gooch crucible to vacuum.
40. Pour centrate into Gooch crucible.

41. Rinse remains in centrifuge tube in Gooch crucible with distilled water.
42. When glass filter is dry turn off vacuum.
43. Disassemble Gooch crucible and stopper, rinse surfaces of these components with distilled water into vacuum flask associated with Gooch crucible.
44. Pour contents of vacuum flask into crucible number 4.
45. Rinse vacuum flask with distilled water into crucible number 4 and place in drying oven.
46. After at least 24 hours, remove crucibles 1 to 4 from oven.
47. Reweigh crucibles.

Calculations and Discussion of Procedure:

In many instances the calculations and procedures are standard as they are dictated by logic. However, in some cases various peculiarities of the sludge make standard evaluations difficult. In such instances procedures were designed which allowed for the properties of the sludge to arrive at the designed dependent variables. A discussion of the calculations and procedures is given in the following paragraphs. The calculations will be outlined with respect to the enumerated experimental steps.

Total Solids of Digester Effluent -

This is a standard method of evaluating percent solids. It necessitates the removal of the liquid fraction of the sample by drying the sample in an oven at 103°C. The ratio of the dry weight to the wet weight of the sample yields percent total solids.

T_1 = tare weight of crucible #1

W_1 = weight added to crucible #1

C_1 = weight of crucible #1 and dried solids

$$\text{Sludge total solids (\%)} = \frac{C_1 - T_1}{W_1} \times 100$$

--Possible sources of error --

1. Obtaining representative sample
2. Determining weight added (accuracy of scale used)

Filter Cake Percent Total Solids -

This method is also fairly standard except for the fact that the entire cake is used in the sample. This is done in order to be able to compute a filter capture ratio with a minimum number of assumptions. Using the entire cake is probably a better method as it eliminates the concern about removing a representative sample from the cake.

T_2 = tare weight of crucible #2

W_2 = weight of crucible #2 plus wet cake

C_2 = weight of crucible #2 plus dry cake

$$\text{Cake solids (\%)} = \frac{C_2 - T_2}{W_2 - T_2} \times 100$$

-- Possible sources of error --

1. Ability to remove all the cake from filter
2. Cake dries out in the period between testing and weighing
3. Since entire cake is used the weight of a large cake (corn stover) forces one to use a less accurate balance which has a larger capacity

Filtrate Total Solids -

This test is very standard and identical to the sludge total solids. Obtaining a representative sample is easy since the particles are small and settle slowly.

T_3 = tare weight of crucible #3

W_3 = weight of filtrate added to crucible #3

C_3 = weight of crucible #3 and dry solids

$$\text{Filtrate total solids (\%)} = \frac{C_3 - T_3}{W_3} \times 100$$

-- Possible sources of error --

1. Determining weight added (accuracy of balance used)

Filtrate Dissolved Solids -

This test is one that had to be designed to fit the sludge. This test is usually performed by passing the filtrate through a fiberglass micropore filter thus filtering out the suspended solids. However, because of the concentration and nature of suspended solids, the glass filter was subject to rapid plugging. To eliminate this problem, a known weight of filtrate was centrifuged first to remove the majority of suspended solids. The centrate was passed through the filter and the filtrate analyzed for dissolved solids. The original weight of the sample is used in the denominator of the ratio since theoretically this is correct. An operational error does exist since the water within the pore volume of the settled suspended solids also contains dissolved solids and is not accounted for. This error is believed to be small and smaller than the error associated with other techniques.

T_4 = tare weight of crucible #4

W_4 = weight filtrate placed in centrifuge tube

C_4 = weight of crucible #4 and dried solids

$$\text{Filtrate dissolved solids (\%)} = \frac{C_4 - T_4}{W_4} \times 100 \quad --$$

-- Possible sources of error --

1. Dissolved solids in water retained by suspended solids centrifuge pellet
2. Determination of weight added to centrifuge tube (accuracy of scale)

Filtrate Suspended Solids -

This value was determined as the difference between the total solids and the dissolved solids.

$$\text{Filtrate suspended solids (\%)} = \text{TS} - \text{DS}$$

-- Possible sources of error --

1. Errors associated with filtrate total and dissolved solids

Solids Capture -

The percent solids capture is the ratio of the weight of solids caught on the filter media to the weight of suspended solids entering the system. Although there are formulas which allow the use of a sample of the filter cake in the calculation, they are subject to error due to assumptions made in the derivations. The method used below eliminates those errors since the whole cake is used in the ratio determination.

T_2 = tare weight of crucible #2

W_2 = weight of crucible #2 and wet cake

$W = W_2 - T_2$ = weight of wet cake

X_1 = cake solids (%)

V = volume of filtrate

X_2 = filtrate suspended solids (%)

ρ = density of filtrate $\approx \rho_{H_2O}$

$$\text{Solids capture (\%)} = \frac{\text{weight of solids caught on filter}}{\text{weight of suspended solids entering system}}$$

$$\text{Solids capture (\%)} = \frac{WX_1}{WX_1 + \rho_{H_2O} X_2 V}$$

-- Possible sources of error --

1. Assumption $\rho = \rho_{H_2O}$
2. Errors in terms X_1 and X_2 disclosed previously

Solids Loading Rate -

The solids loading rate is an important design parameter used in sizing vacuum filtration facilities. The units of the solids loading rate is commonly lb dry solids/ft²-hr. The relationship will be derived below:

T_2 = tare weight of crucible #2

W_2 = weight of crucible #2 and wet cake

$W = W_2 - T_2$ = weight of wet cake

A = area of test leaf = 0.1 ft²

FT = form time (s)

DT = dry time (s)

$T = FT + DT$ (s)

X_1 = % cake solids

$$\text{Solids loading rate} = \frac{WX_1}{0.1 \text{ ft}^2 T} \times \frac{\text{LB}}{453.9 \text{ g}} \times \frac{3600 \text{ S}}{\text{hr}} = 79.3 \frac{WX_1}{T}$$

To account to dead time the above figure is usually multiplied by a factor of 0.8 (Eimco, 1975).

$$\text{Solids loading rate} = 63.44 \frac{WX_1}{T} \frac{\text{LB}}{\text{ft}^2\text{-hr}}$$

4. Routine Analyses

Routine tests for control of the fermentation process and for evaluation of the performance were conducted as outlined by Standard Methods (1976).

APPENDIX C

COMPUTER PROGRAM AND DOCUMENTATION

This appendix contains detailed information on the simulation program. It is divided into four major sections: logic flow and notes, definition of major program variables, program listing, and sample output. The program has been implemented on the University of Illinois CDC CYBER-175 computer system in a time-sharing mode.

1. Logic Flow and Notes

1.1 Overview of Simulation Program

The overall structure of the system being simulated is shown in Figure C-1. The computer program consists of separate subroutines for each of the operations shown in the figure, with additional output and data interpolation subroutines as required. Major emphasis has been placed on the construction of a reasonably accurate model of the anaerobic digestion process, with less importance attached to the system components before and after digestion.

Major portions of this simulation program are alterations of earlier models of beef lot manure digestion (Pfeffer and Quindry, 1978) and organic refuse digestion (Quindry, Liebman and Pfeffer, 1976). Other portions are based on a manure digestion program developed by Dynatech R/D Company (Ashare, Wise and Wentworth, 1977).

1.2 Data Input and Initialization

Data input for the simulation program consists of both a file of parameters (named FRONT) and free-format interactive data that is input during execution of the program. Many parameters which would be changed

Figure C-1. Overall Structure of Computer Program

only rarely are initialized in DATA statements at the beginning of the subroutine in which the parameters are used.

File FRONT consists of information which can be changed more readily and includes such items as interest rate, inflation adjustments for the capital cost items, mixing horsepower requirements, electricity cost and methane selling price.

Selection of system options, reactor temperature, heat exchanger temperature gradient, and many other parameters is accomplished through the prompting by the simulation program for the desired information when it is needed. These parameters can be changed for each simulation run, and several simulation runs can be done with a single execution of the program.

1.3 Predigestion Systems: Caustic Treatment, Mixing, and Heat Exchange

The raw input materials available as agricultural residue are not suitable for anaerobic digestion without initial processing. In addition to those measures which are required to permit anaerobic digestion to take place, system performance and economics may be improved by the inclusion of other predigestion processes.

The initial input material is assumed to be a solid material with relatively low moisture content and with negligible nutrient content. Since the mixing of the material and its physical transport through the digestion process requires a moisture content of around 8 percent, considerable water must be added and mixed with the material to be digested. The efficiency of the digestion process is greatly affected by digester operating parameters such as temperature, nutrient level, pH, etc. These operating parameters can be affected by predigestion processing. Specifically, heat treatment with caustic affects digester pH and temperature, as well as increasing the

digestibility of the substrate. If caustic treatment is not employed, some nutrient and heat recovery can be accomplished through recycle of digester effluent after removing most of the suspended solids by centrifuge. The combination of recycle and heat treatment is not feasible due to a buildup of sodium in the system. As another option, heat recovery alone may be accomplished through the use of a heat exchanger. Each of these options is included as an optional subroutine in the digestion program.

Subroutine MIX includes the material and energy balances for these predigestion systems, and the sizing and cost calculations for the necessary equipment. The slurry solids concentration is a variable input to the system and the capacity of the mix tank, which provides storage capacity to smooth digester inflow rate, can be changed by altering the value of the mix tank retention time (currently set at 1 hour).

1.4 Anaerobic Digester Model

The important parameters for the digestion system are temperature, retention time and slurry concentration, and the rate of input of biodegradable solids, alkalinity and nutrients in the feed stream. From the retention time, concentration and solids input rate, the volume required for digestion can be calculated (line 41500). This volume is then divided into the number of digesters required according to the data in Table C-1.

1.4.1 Volatile Solids Destruction and Gas Production

The total amount of CO_2 and CH_4 produced by the bacteria is calculated using the digestion temperature, retention time and volume, and the biodegradability factor and kinetic constant as shown on pages 25 through 27 of this report. All of the carbon dioxide produced by the fermentation of the substrate is not discharged in the gas phase. Because of the

Table C-1
Number of Digesters Required

Total Volume (cubic meters)	Number of Digesters
0 to 3,682	2
3,682 to 14,160	3
14,160 to 35,400	4
35,400 to 48,144	5
48,144 to 56,640	6
more than 56,640	7

relatively high solubility of carbon dioxide, and its reaction with water to produce carbonic acid, a portion of the CO_2 is removed from the digester in the liquid effluent. The quantity of CO_2 removed in this manner depends on the digestion temperature, the rate of input of alkalinity and nitrogen from the input slurry and digester pH. Digestion temperature affects the dissociation constant of carbonic acid (K_1), the dissociation constant of ammonia (K_N), Henry's Constant (H_K) and the vapor pressure of water (V_p) as shown in Table C-2.

Table C-2
Effect of Temperature on Digester Constants

Constant	58°C
K_1	5.178×10^{-7}
K_N	6.136×10^{-9}
H_K	2.5×10^6
V_p (atmos.)	0.1790

Source: Handbook of Chemistry and Physics, 48th ed., CRC Publishing Co.

The actual rate of gaseous CO_2 production can be estimated by calculating the equilibrium conditions in the digester, and subtracting the soluble CO_2 and associated bicarbonate from the total CO_2 input. The reaction of CO_2 in water to produce carbonic acid is given as:

Carbonic acid dissociates readily to form a hydrogen ion and bicarbonate:

The degree of dissociation is dependent on temperature and pH,

$$K_1 = [\text{H}^+] \times [\text{HCO}_3^-] / [\text{H}_2\text{CO}_3] \quad (\text{C-3})$$

where K_1 is taken from Table C-2. The reaction between carbon dioxide and water is essentially complete and the concentration of carbon dioxide in solution as assumed equal to the concentration of carbonic acid.

Alkalinity is added to the digester from sources other than carbon dioxide production. Other sources that must be included are the dilution water added with the substrate, the caustic pretreatment of the substrate, the recycled water from the dewatering step, and the alkalinity of the substrate itself. Calculations in the program are based on the estimated alkalinities given in Table C-3.

The alkalinity contribution (in moles per liter of digester per day) is calculated by dividing the values in Table C-3 by 100 (to convert from grams as CaCO_3 to moles) and dividing by the total digester volume.

The amount of CO_2 in the liquid phase is also affected by the ammonia concentration in the input. Ammonia reacts with water to produce

Table C-3
Estimated Alkalinities for Input Streams

Material	Alkalinity - g/t
Dilution Water	0.1
Recycled Centrate	6.0
Corn Stover	
Without Caustic Treatment	0.0
With Caustic Treatment	4.0

NH_4^+ and OH^- .

The production of hydroxide affects the pH of the digester, which in turn alters the CO_2 solubility. For the purpose of calculating the equilibrium conditions in the digester, each mole of hydroxide adds an equivalent of one mole of alkalinity to the digester (line 45900).

The minimum digester pH is assumed to be 6.8, with this level being maintained by the addition of lime or sodium hydroxide if required. With caustic pretreatment, a digester pH higher than 6.8 may be maintained without the addition of additional caustic to the digesters. In this case, the pH must be calculated so that the quantity of gaseous CO_2 can be determined.

If the digester pH is known, it is possible to calculate all the remaining unknowns by solving a quadratic equation (lines 46000 to 47100 in the program). The procedure used to calculate the equilibrium pH is to assume a very low pH, do all the required calculations for

determining ammonium ion, bicarbonate alkalinity, gaseous CO_2 , and soluble CO_2 concentrations (lines 45000 to 47000) and to then calculate the amount of hydroxide ion addition that is required to maintain the assumed pH. For the low pH that is initially assumed this value will be negative, indicating that the pH should be raised. The pH is then incremented upward and the calculations are done again. This sequence is repeated until the hydroxide requirement becomes slightly positive. Once the equilibrium concentrations for all the end products are known, it is possible to calculate volatile solids destruction, volatile solids remaining, effluent slurry concentration and the methane content of the gas stream (lines 47800 to 48200).

1.4.2 Cost Calculations

The total cost of the digester system depends on both the capital cost and the operating cost. Capital cost is calculated from data presented by Patterson and Banker (1971). The costs given in that publication have been fitted to equations as shown in Table C-4 (lines 48300 to 49300 of the program).

Table C-4
Capital Cost for Digesters

Digester Volume (TVD) (meters ³)			Capital Cost (1971 dollars)
0	TVD	1416	\$165,000
1416	TVD	2832	$45.9 \times \text{TVD} + 100,000$
2832	TVD	8496	$51.2 \times \text{TVD} + 85,000$
	TVD	> 8496	$1000 \times 10^{[0.987 \log (0.0531 \text{ TVD}) + 0.271]}$

Annual maintenance cost is assumed to be one percent of the installed capital cost (line 49400). Mixing costs are assumed to be linear with volume and solids concentration. An empirically chosen energy requirement (0.475 horsepower per 1000 cubic feet at 4 percent solids) is multiplied by the existing solids concentration and volume ratios. The value of the mixing energy requirement is then converted to the metric system.

1.4.3 Energy Requirements

A significant amount of energy is required to maintain digester temperature at the desired level. This requirement is the sum of three components:

a) Heatloss through digester walls

Three separate equations are used, for the sides, top and bottom. Each is of the form:

$$HL = \frac{1.8 (A)(T-T_0)}{(1/FI + B/K + 1/F0)} \quad (C-5)$$

where

HL = heatloss - BTU/hr

A = area - square feet

T = digester temperature - °C

T₀ = ambient temperature - °C

FI = inside transmittance coefficient

F0 = outside transmittance coefficient

B = length of heat flow path - inches

K = thermal conductivity coefficient for wall material

After calculation of the heatloss in BTU/hr (lines 51000 to 51500) the value is converted to Joules/hr (line 53810).

b) Heat required to increase the temperature of the incoming slurry is calculated using the output temperature of the caustic treatment tanks, the heat exchanger, or the ambient temperature, depending on the pretreatment process employed. If centrate recycle is included in the simulation, the recycled water is assumed to be at 50°C, with an assumed loss of 10°C through the dewatering phase.

c) Heatloss in water vapor is calculated by using an empirical equation developed for the refuse digestion program (Quindry et al., 1976).

1.5 Boiler System

In order to provide the heat required for the digesters and gas cleanup, a boiler system using raw digester gas is included. The heat requirement for the digester is known from the digester subroutine. If a gas cleanup system is to be included, the energy required to regenerate the monoethanolamine (MEA) used in the regeneration process is estimated, and the amount of gas containing this total energy requirement is removed from the gas stream. The boiler system is sized to produce the required energy output, and cost figures are calculated from available estimates (Blecker and Nichols, 1973).

1.6 Gas Cleanup

The optional gas cleanup system is intended to remove both the large quantity of carbon dioxide and the relatively small quantities of hydrogen sulfide that may be present in the gas stream. If on-site use of the gas is contemplated, rather than transmission of the gas through a pipeline, carbon dioxide removal would not be included.

1.7 Slurry Dewatering by Centrifugation

Two limiting conditions exist for the performance of centrifugation. One is the slurry flow rate and the other is the solids loading. The number of centrifuges depends upon which of the two limiting conditions dominates for a given operating scheme. The solids loading on the centrifuge relates to cake dryness. Experimental results with a fibrous material, urban refuse, were used in sizing the centrifuge. The maximum slurry feed rate was found to be $6.37 \text{ m}^3/\text{min}$ (225 gpm). This limiting condition will apply at low slurry solids concentration. However, the solids loading will generally apply when the slurry solids concentration exceeds 3 percent. The determination of the number of machines required (lines 70000 to 70380) is then used to calculate the costs of centrifuge and the output of the centrifuge (lines 70400 - 71380). Capital cost includes machine cost, installation and building. Based on current information, installation costs are assumed to be 50 percent of the total machine cost.

2. Definition of Major Program Variables

A(1)	Roof surface area of digester
A(2)	Wall surface area of digester
A(3)	Bottom surface area of digester
AANBIO	Percent biodegradable solids
AHEAT	Area of heat exchanger
AJUSE	Energy used (Joules)
AK	Rate constant day^{-1}
AKN	Ammonia dissociation constant
AKTHS	Substrate utilization
ALATHT	Latent heat vaporization of water

ALK	Alkalinity (g/l as CaCO_3)
ALKC	Alkalinity of substrate after caustic treatment
ALKD	Alkalinity of dilution water
ALKM	Alkalinity of substrate
ALKREC	Alkalinity of recycled centrate
ALR	Liquid limiting flow rate for centrifuge
AMCAP	Amortized capital
AMCR	Maintenance cost/capital cost ratio
AMM	Ammonia input with substrate
AMON	Ammonia concentration in digester
ANH3	NH_3 concentration (moles)
ANH3N	NH_3 concentration (mg/l as N)
ANH4	NH_4^+ concentration (moles)
APCOST	Amortized pump cost
ASH	Ash in feed stream (kg/hr)
B	Digester wall thickness (inches)
BICO	Bicarbonate alkalinity (moles/l)
BIOD	Biodegradable solids flow rate (kg/hr)
BOIOP	Boiler operation and maintenance cost
BTUUSE	Boiler capacity (BTU/hr)
CAKE	Total centrifuge solids discharged
CAKEMO	Centrifuge cake moisture
CAKETS	Centrifuge cake solids
CAKEYS	Centrifuge cake volatile solids
CAP	Total capital required
CAUADD	Caustic feed rate (lbs/lb substrate)
CAUOMC	Caustic treatment O & M cost

CAUS	Cost of NaOH (\$/hr)
CCAKE	Centrifuge cake solids concentration (percent)
CCBLD	Centrifuge building capital cost
CCBOI	Boiler capital cost
CCBOIA	Amortized boiler capital cost
CCEN	Centrate solids concentration (percent)
CCENT	Capital cost of centrifuges
CCINST	Centrifuge installation costs
CCXCH	Heat exchanger capital costs
CCXCHA	Amortized heat exchanger capital costs
CCWAT	Capital cost of boiler water treatment system
CCWATA	Amortized cost of boiler water treatment system
CEN	Total centrate
CENCCA	Amortized centrifuge capital cost
CENMC	Maintenance cost for centrifuges
CENMO	Centrate moisture
CENTS	Total centrate solids
CH4	Methane production (m^3/hr)
CH4BEN	Credit for methane (\$/hr)
CH4BTU	Heat value of methane (Btu/hr)
CH4CH	Same as CH4
CH4HOU	Heat value of methane (Joules/hr)
CONC	Digester solids concentration (percent)
CO2	Carbon dioxide in gas phase - final value (m^3/hr)
CO2B	Carbon dioxide added to digester by fermentation (moles%/day)
CO2CH	Gaseous carbon dioxide flow rate
CO2GCH	Same as CO2CH

C02P	Total carbon dioxide produced
C02S	Dissolved carbon dioxide
CRR	Capital recovery factor
C4	Mixing power requirement at 4 percent solids
C5	Conversion factor HP to KW
C6	Cost of electricity (\$/KWH)
C9	Methane value (\$/million BTU)
C10	Conversion factor HP to BTU/hr
C16	Constant for centrifuge loading calculation
C17	Constant for centrifuge loading calculation
DAMIX	Retention time in mix tank (days)
DD	Digester diameter
DH	Digester height
DIGCON	Solids concentration out of digester
DIGMC	Digester maintenance cost
DLIFE	Design life for system components
DPROF	Value of recovered gas (\$/hr)
ENREM	Energy content of feed stream (BTU/hr)
ENREMJ	Energy content of feed stream (Joules/hr)
ENREQ	Energy required for system operation (BTU/hr)
ENREQJ	Energy required for system operation (Joules/hr)
FI	Heat transmittance coefficient, inside
FO	Heat transmittance coefficient, outside
FRAC	Methane fraction of gas stream
GASCH	Total gas produced (ft ³ /hr)
GAST	Total gas produced (m ³ /hr)

GCLIFE	Gas cleanup system design life
HCDST	Hourly cost of individual processes
HCO3D	Bicarbonate alkalinity added with dilution water
HCO3M	Bicarbonate alkalinity added with manure
HK	Henry's Constant
HL	Heat loss in digester
HLT	Total heat loss in digester
HR	Heat required for digester (BTU/hr)
HRJ	Heat required for digester (Joules/hr)
HVAL	Heat value of gas stream (BTU/hr)
HVALJ	Heat value of gas stream (Joules/hr)
HVGC	Heat required for gas cleanup (BTU/hr)
HVG CJ	Heat required for gas cleanup (Joules/hr)
H2O	Water in gas stream
H2OFR	Water fraction of gas stream
K	Conductance coefficient for digester
MEACC	Gas cleanup capital cost
MEACCA	Amortized gas cleanup capital cost
MEADMC	Gas cleanup operating and maintenance costs
MIXBTU	Mixing energy required
MIXCC	Mix tank capital cost
MIXCCA	Amortized mix tank capital cost
MIXHP	Mix tank motor size
MIXOMC	Mix tank operation and maintenance cost
MIXVOL	Mix tank volume
MUMEA	Make-up MEA
MUMEAC	Make-up MEA cost

MWMEA	Molecular weight of MEA
NUMBR	Substrate input rate (tons/day)
NU	Number of digesters required
OANDM	Total cost of operation and maintenance
OH	Hydroxide requirement
OMCCR	Operation and maintenance/capital cost ratio
PBTUR	Pumping energy requirement
PC	Centrifuge power cost
PCOST	Pumping capital cost
PH	Digester pH
PHP	Pump horsepower
POMC	Pumping operation and maintenance cost
QLIMT	Centrate limiting flow rate
RATIO	Digester height/radius ratio
RD	Digester radius
RET	Digester retention time (days)
RINST	Installation cost/machine cost for centrifuge
SCOUT	Solids concentration in digester effluent (kg/kg)
SLTEMP	Slurry temperature (°C)
SMEAC	MEA circulation rate
SQ1	Initial substrate level
SPHWV	Specific heat of water vapor
SR	Solids flow rate in each centrifuge
SRATE	Limiting solids flow rate
STCI	Current inflation index
STCmyy	Inflation index for month m of year yy
STEAM	Gas clean-up steam requirement

STEP	Character array
TBTUC	Total energy consumed
TCAPT	Total capital cost for centrifugation
TCCD	Total capital cost for digesters
TCCDA	Amortized digester capital cost
TDIF	Heat exchanger temperature gradient ($^{\circ}\text{C}$)
TEMP	Slurry temperature ($^{\circ}\text{C}$)
THLD	Total heat loss in digesters (BTU/hr)
THR	Total heat required (BTU/hr)
THRJ	Total heat required (Joules/hr)
TO	Temperature of incoming slurry to heat exchanger
TOTVOL	Total mix tank volume
TS	Total solids in slurry
TSREM	Total solids in slurry after digestion
TUD	Total digester volume
VAREA	Unit centrifuge area
UCBLD	Unit cost of centrifuge building
UCCENT	Unit cost of centrifuge
UP	Power requirement for unit centrifuge
VOLD	Volume of each digester
VP	Vapor pressure of water
VS	Volatile solids in slurry (kg/hr)
VSD	Volatile solids destroyed (kg/hr)
USREM	Volatile solids in slurry after digestion (kg/hr)
WATIN	Water input in substrate (L/hr)
WATNED	Make-up water required (L/hr)
WATR	Total water input to system (L/hr)

XINT	Interest rate (fraction)
XINTP	Interest rate (percent)
X1	Quadratic equation solution
X2	Quadratic equation solution

3. Program Listing

```

00100      PROGRAM CCRNST (INPUT, OUTPUT, FRONT, POUT, TAP15=INPUT, TAP26=OUTPUT,
00200      1 TAPE9=FRONT, TAPE13=POUT)
00300      COMMON/UCOM1/WATR,TS,VS,BIOD,ASH,AMH,ALK,PH,DIGCON,VOL0,CH4,C02,
00400      2 H2O,
00500      1 HVAL,CAP,ANCAP,OADNH,DPROF,ENREU,ENREN,WATNEU,WATIN,INCP
00600      COMMON/UCOM2/XINT,STC767,STC772,STC773,STC174,STC175,STC775,STC1,
00700      2 STC771,
00800      1 C4,C5,C6,C9,C10,S01,AK1,RET,CCNC,TEMP
00900      2 WRITE(6,5)
01000      5 FORMAT(1X,"RESIDUE DIGESTION PROGRAM"/)
01100      PRINT(6,*) "CAPACITY--TPD",
01200      READ(5,*) NUMBR
01300      PRINT(6,*) "CONCENTRATION INPUT TO DIGESTER (%)",
01400      READ(5,*) CONC
01500      PRINT(6,*) "DESIRED DAYS RETENTION",
01600      READ(5,*) RET
01700      PRINT(6,*) "DESIRED RECYCLE RATE",
01800      READ(5,*) RECY
01900      WRITE(3,9) NUMBR, CONC, RET, RECY
02000      9 FORMAT("1", "RESIDUE DIGESTION SIMULATION"///
02100      14X, "PLANT CAPACITY.....", I12, " TONS/DAY"/,
02200      24X, "DIGESTER INPUT CONCENTRATION.....", F12, P5.2, " PERCENT"/,
02300      44X, "RETENTION TIME REQUESTED.....", F12, P5.2, " DAYS"/,
02400      54X, "RECYCLE RATE FOR CENIRATE.....", F12, P5.2, " PERCENT"/)
02500      12 CONTINUE
02600      PRINT (3,*) " "
02700      READ(9,*) XINT,STC767,STC771,STC172,STC772,STC773,STC174,STC175,
02800      1 STC775,STC1,C4,C5,C6,C9,C10
02900      CALL INITAL(NUMBR)
03000      CALL PRNT(1)
03100      CALL MIX(NUMBR,RECY,SLTEMP)
03200      CALL PRNT(2)
03300      CALL DIGES(RECY,SLTEMP,THR)
03400      CALL PRNT(3)
03500      CALL HEAT(THR,IGC)
03600      CALL PRNT(6)
03700      IF(IGC.NE.1) GO TO *5
03800      CALL CLEAN
03900      CALL PRNT(5)
04000      15 CONTINUE
04100      PRINT(5,*) "INCLUDE DEWATERING 1) YES, 2) NO",
04200      READ(6,*) IDW
04300      IF(IDW.NE.1) GO TO 17
04400      33 CALL CENTR
04500      CALL PRNT(4)
04600      17 CALL FINAL
04700      PRINT(6,*) "1) REPEAT ANALYSIS OR 2) STOP",
04800      READ(5,*) ISTOP
04900      IF(ISTOP.EQ.1) GO TO 20
05000      STOP
05100      20 REWIND 9
05200      GO TO 2
05300      END
05400      SUBROUTINE INITAL(NUMBR)
05500      COMMON/UCOM1/WATR,TS,VS,BIOD,ASH,AMH,ALK,PH,DIGCON,VOL0,CH4,C02,
05600      2 H2O,
05700      1 HVAL,CAP,ANCAP,OADNH,DPROF,ENREU,ENREN,WATNEU,WATIN,INCP
05800      COMMON/UCOM2/XINT,STC767,STC772,STC773,STC174,STC175,STC775,
05900      1 STC1,STC771,C4,C5,C6,C9,C10,S01,AK1,RET,CCNC,TEMP
06000      DIMENSION CHAR(2,4),AK(2)
06100      DATA CHAR/15.,15.,85.,85.,90.,90.,40.,77./

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11050 DATA VSBTU/8600.0/
11100 DATA AK/.25,.25/
11110 PRINT(6,*) "INCLUDE CAUSTIC PRETREATMENT? 0-NO, 1-YES ",
11120 READ(5,*) INCP
11130 LTYPE=INCP+1
11200 IF (CONC.LE.CHAR(LTYPE,2)) GO TO 100
11300 WRITE(6,10) CONC,CHAR(LTYPE,2)
11400 10 FORMAT(1X,"INPUT CONCENTRATION OF ",F5.2," PERCENT IS ABOVE "
11500 1"THE"/,"X,"SUBSTRATE SOLIDS CONCENTRATION OF"F6.2" PERCENT")
11600 STOP
11700 100 CONTINUE
11800 C ****SET VARIABLES TO INITIAL VALUES
12800 ENREQ=0.
13000 CAP=0.
13100 AMCAP=0.
13200 OANDN=0.
13300 CH4=0.
13400 CO2=0.
13500 H2O=0.
13600 ALK=0.
13700 PH=0.
13800 VCLD=0.
13900 DIGCON=0.
14000 HVAL=0.
14100 DPROP=0.
14250 AK1=AK(1)
14327 71 WRITE(6,72) AK1, (CHAR(LTYPE,I), I=1,4)
14329 72 FORMAT("1PARAMETERS USED FOR SOLIDS DESTRUCTION"//,
14335 34X," K.....",8X,F4.2," PER DAY"//,
14337 44X,"COMPOSITION OF FEED MATERIAL"/,
14339 54X," WATER.....",7X,F5.2," PERCENT"/,
14341 64X," TOTAL SOLIDS.....",7X,F5.2," PERCENT"/,
14343 74X," VOLATILE SOLIDS.....",7X,F5.2," PERCENT T.S."/,
14345 84X," BIODEGRADABLE SOLIDS.....",7X,F5.2," PERCENT V. S."/)
14349 PRINT(6,*) "DO YOU WANT TO CHANGE ANY OF THESE VALUES"
14351 PRINT(6,*) "1-YES 0-NO",
14353 READ(5,*) IANS
14355 IF (IANS.EQ.0) GO TO 74
14357 PRINT(6,*) "INPUT ALL 6 VALUES IN THE ABOVE ORDER"
14359 READ(5,*) AK1, (CHAR(LTYPE,I), I=1,4)
14361 74 WRITE(3,72) AK1, (CHAR(LTYPE,I), I=1,4)
14400 AAN=CHAR(LTYPE,1)
14500 AANTS=CHAR(LTYPE,2)
14600 AANVS=CHAR(LTYPE,3)
14700 AANBIO=CHAR(LTYPE,4)
14805 SO1=.63*CHAR(LTYPE,4)/100.
14810 ENREQ=NUMBER*2000.*VSBTU*AANVS*AANTS/24./10000.
14820 TS=NUMBER*AANTS*.37793
14830 VS=TS*AANVS/100.
14835 ASH=TS-VS
14840 BIOD=AANBIO/100.*VS
14850 WATR=TS/(AANTS/100.)-TS
14860 AMN=.00
14870 IF (INCP.EQ.0) RETURN
14880 CALL CAUST
14900 RETURN
15000 END
20000 SUBROUTINE FRNT(I)
20100 CCONH/UCON1/WATR,TS,VS,BIOD,ASH,AMN,ALK,PH,DIGCON,VOLJ,CH4,CO2,
20150 2 H2O,
20200 1 HVAL,CAP,AMCAP,OANDN,DPROP,ENREQ,ENREQ,WATBED,WATID,INCP

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[illegible]

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32200      WRITE(3,10) CCNC,WATNED
32300      10 FORMAT(4X,"TOTAL WATER REQUIREMENTS FROM RECYCLE AND"/,
32400      14X,"MAKEUP FOR ",F5.2," PERCENT INPUT SOLIDS:"/,
32500      24X,G12.5," L/HR")
32600      ENGL=WATNED/3.78
32700      WRITE(3,11) ENGL
32800      11 FORMAT(46X,G12.5," GAL/HR"///)
32900      SLTEMP=10.
32950      RECMAT=0.
32960      AMNR=.25E-03
33000      IF(RECY.LF.0) GO TO 100
33100 C ****ADD WATER, HEAT FROM RECYCLE
33110      REMR=.1E-03
33200      RECMAT=RECY/100.*WATNED
33300      SLTEMP=SLTEMP+(TEMP-20.)*RECMAT/(WATOUT+TS)
33400 C ****CALCULATE MIXING EQUIPMENT SIZE
33500 C ****ASSUME 1 HOUR HOLDING IN MIX TANK
33600 C ****VOLUMES IN LITERS
33700      100 TOTVOL=(WATOUT+TS)*24.
33800      DAMIX=1./24.
33900      MIXVOL=DAMIX*TOTVOL
34000 C ****CALCULATE MIXER CAPITAL COST
34100 C ****REFERENCE DYNATECH, PG. 167
34200      MIXCC=STCI/STC771*13722*(MIXVOL/28316.)**.52
34250      PRINT(6,*)MIXCC,STCI,STC771,MIXVOL
34300 C ****AMORTIZE CAPITAL COST OVER 20 YEARS
34400      MIXCCA=MIXCC/360./24.*XINT/(1.-(1.+XINT)**(-20))
34500 C ****OP. AND MAINTENANCE IS A FRACTION OF AMORTIZED CAPITAL COST
34600      MIXOMC=OMCCR*MIXCCA
34700 C ****ASSUME .2 HP REQUIRED PER MCF FOR MIXING
34800 C ****REFERENCE DYNATECH
34900      MIXHP=0.2*MIXVOL/28320.
35000 C ****CONVERT TO BTU/HR
35100      MIXBTU=MIXHP*C10
35200 C ****GET PUMPING POWER REQUIREMENT BY SCALING UP OR DOWN
35300 C ****REFERENCE REFUSE DIGESTER PROGRAM
35400      PHPREQ=PHP*IS/20416.66
35500      PPTUR=PHPREQ*C10
35600 C ****PUMP COSTS ASSUME $75./HP AT 1973 COSTS
35700      PCOST=75.*PHPREQ*STCI/STC773
35800      AECOST=PCOST/360./24.*XINT/(1.-(1.+XINT)**(-20))
35900      PCMC=OMCCR*AECOST
35910      AMNR=AMNR+WATOUT
35920      APHOS=AMNR*.2
35930      CNUTR=AMNR*.15*2.2+APHOS*.15*2.2
35940      WRITE(3,217)CNUTR
35950      217 FORMAT("NUTRIENT COST.....",F12.2," $"///)
35960      AMH=.25E-03*WATR
36000 C ****ADD TO TOTALS FOR PRINT SUBROUTINE
36100      CAP=CAP+MIXCC+PCOST
36200      AMCAP=AMCAP+MIXCCA+AECOST
36300      OANDM=OANDM+MIXOMC+PCMC+CNUTR
36400      WATR=WATOUT
36500      ERREQ=ENREQ+PPTUR+MIXBTU
36600      ASH=TS-VS
36900      PRINT(6,*)"ADD HEAT EXCHANGER? 1=YES, 0=NO",
37000      READ(5,*)IAND
37100      IF(IAND.NE.1) RETURN
37125      PRINT(6,*)"TEMPERATURE GRADIENT FOR EXCHANGER (DEG. C)",
37150      READ(5,*)TDIF
37200      WRITE(3,78)

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37300      70 FORMAT(1X,"HEAT EXCHANGE PARAMETERS"/)
37400      AHEAT=.00195*(WATR+TS)*24.*(TEMP-SLTEMP-TDIF)/TDIF
37500      CCXCH=26.*SICI/STC767*(AHEAT)**.75
37600 C ****CONVERT TO SQUARE METERS
37700      AHEAT=AHEAT*.0929
37800      CCXCHA=CCXCH/360./24.*(KINT/(1.-(1.+KINT)**(-20.)))
37900      KCHOMC=OMCCF*CCXCHA
38000      WRITE(3,79) AHEAT,TDIF,CCXCH,CCXCHA,XCHOMC
38100      79 FORMAT(4X,"SURFACE AREA....." G12.5," Sq. M."/
38150      44X,"TEMPERATURE GRADIENT....." 7X,F5.1," DEG. C"/,
38200      14X,"CAPITAL COST....." F12.2," $"/
38300      24X,"AMORTIZED CAPITAL COST....." F12.2," $/hr"/,
38400      34X,"OPERATION AND MAINTENANCE....." F12.2," $/hr"/)
38500      CAP=CAP+CCXCH
38600      ANCAP=ANCAP+CCXCHA
38700      SLTEMP=TEMP-TDIF
38800      CANDM=OANDM+XCHOMC
38900      RETURN
39000      END
40000      SUBROUTINE DISES(RLCY,SLTEMP,TH2)
40100      REAL MP,NEZ,dPC,K,K1,qch
40200      COMMON/UCCH*/WATR,TS,VS,BIOD,ASH,AMH,ALK,PH,SCOUT,TVJ,CH4CH,
40300      1 CO2GCH,U20,HVAL,CAP,ANCAP,OANDM,OPDOF,KNREG,FMLED,ATMED,
40400      1 WATR,INCF
40500      COMMON/UCCH2/KINT,STC767,STC772,STC773,STC774,STC775,STC776,
40600      1 STC777,C4,C5,C6,C9,C10,S01,AK1,RET,CCMC,TEMP
40800      DIMENSION VR(5),TO(3),A(3),HL(3),FI(3),B(3),K(3),FO(3)
40900      IPTA FI,B/3*2.,8.,18.,12./
41000      DATA K,FO,B/3*8.3,2*1.3,2.,2*12.5,15./
41100      DATA NVDATA,VR/5,3681.0,14160.,35400.,48140.,56640./
41110      DATA TEMP/58./
41200      DATA TO/12.5,12.5,15./
41400 C ****VOLUME OF DIGESTER(S) AND NUMBER
41500      TVD=(WATR+TS)*24.*RET/1000.
41600      IF(TVD.LT.VR(1)) GO TO 1015
41700      IF(TVD.GE.VR(NVDATA)) GO TO 1017
41800      GO TO 1020
41900      1015 NV=2
42000      GO TO 1026
42100      1017 NV=7
42200      GO TO 1026
42300      1020 DC 1025 N=2,NVDATA
42400      IF(TVD.LT.VR(N)) GO TO 1023
42500      GO TO 1025
42600      1023 NV=N+1
42700      GO TO 1026
42800      1025 CONTINUE
42900      1026 VI=TVD/NV
43000 C ****CALCULATE VOLATILE SOLIDS DESTRUCTION AND GAS PRODUCTION
43100      SCUT=S01/(AK1*RET+1.)
43200      AKINS=AK1*RET*SCUT
43300      GAST=AKINS*BIOD
43400      CH4CH=GAST*.6
43500      CO2P=CH4CH
43600 C ****CARBONATE EQUILIBRIUM CALCULATIONS
43700 C ****ASSUME A PH AND CALCULATE CONDITIONS
43800      PH=6.8
43900      CC2B=CO2P/TVD*1.071
44000      HK=.139E07
44100      AK=5.04E-07
44200      VF=.0728

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44300 IF (TEMP.LT.50) GO TO 700
44400 HK=.25E07
44500 AK=5.178E-07
44600 VF=.179
44700 700 CCNTIME2
44800 ALKEEC=6.
44900 NIKD=0.1*(1.-ELCY/100.)*ALKEEC*(ELCY/100.)
44925 ALKN=0.01
44950 IF (INCP.EC.7) ALKN=4.
44975 HCO3H=(ALKN/100.)*WATIM*24./(TVD*1000.)
45000 HCO3D=(ALKD/100.)*WATHEL*24./(TVD*1000.)
45100 ANOH=ANH*24./(IVD*14.)
45200 AKN=1.572-05
45300 IF (TEMP.LT.50.) GO TO 701
45400 AKN=6.136E-09
45500 701 CCNTIME2
45600 ANH4=(ANOH*10.**(-PH)/AKN)/(1.+10.**(-PH)/AKN)
45700 ANH3=ANOH-ANH4
45800 ANH3N=ANH3*14000.*RET
45900 BICO=ANH4+HCO3B+HCO3N
46000 AA=18.*HK*10.**(-PH)*RET/(7.6E05*(10.**(-PH)+AK))
46100 BB=1.-VP-AA*BICO/RET
46200 CC=(-AA)*(CC2H**2+CO2B*BICO/RET)
46300 LL=SQR7(BB**2-4.*A*CC)
46400 X1=(-BB+DD)/(2.*AA)
46500 X2=(-BB-DD)/(2.*AA)
46600 IF (71.GT.X2) GO TO 865
46700 CC2G=X2
46800 GO TO 870
46900 865 CC2G=X1
47000 870 CO2S=7.6E05*CO2G*(1.-VP)/(18.*HK*(CO2B+CO2G))
47100 OR=CO2B-CO2G+0.2/(60.*RET)-CO2S/RET-BICO
47200 IF (OR.GE.0.) GO TO 871
47300 PH=PH+.05
47400 IF (PH.GT.8.5) STOP
47500 WRITE(6,27) PH
47600 27 FORMAT(1X,"PH INCREASED TO ",F5.2)
47700 GO TO 701
47800 871 CC2GCH=CO2G+TVD*22.4/24.
47900 GASCH=CO2GCH+CH4CH
48000 VSD=CH4CH*44.64*.162/3.
48010 BIOD=BIOD-VSE
48050 ALK=BICO
48100 VSEEM=VS-VSD
48200 TSEEM=TS-VSE
48300 IF (TVD.GT.1416.) GO TO 910
48400 TCCD=165000.*STCI/STC771
48500 GO TO 917
48600 910 IF (TVD.LT.2832.) GO TO 914
48700 IF (TVD.LT.8496.) GO TO 916
48800 TCCD=10.**(.987*A LOG10(TVD*35.31/1000.)+.271)*1000.*STCI/STC771
48900 GO TO 917
49000 914 TCCD=(1.3*TVD*35.31/1000.+100.)*1000.*STCI/STC771
49100 GO TO 917
49200 916 TCCD=(1.45*TVD*35.31/1000.+85.)*1000.*STCI/STC771
49300 917 TCCDA=TCCD/360./24.*XINT/(1.-(1.+XINT)**(-20.))
49400 DIGAC=0.01*TCCD/24./360.
49500 CH4BTU=CH4CH*35.31*1000.
49600 CH4BEY=Cd4BTU/1.0E+06*C9
49700 SCOUT=TSEEM/(TSEEM+WATR)*100.
49800 HP=0.475*SCOUT/4.00*TVD/1000.*35.31

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49900 MEZ=C10*MP
50000 TFC=NV*C5*C6
50100 DH=48.
50200 910 A(1)=VD*35.31/DH
50300 A(3)=A(1)
50400 BC=SQR1(A(1)/3.14)
50500 RATIO=DH/BD
50600 IF(RATIO.LT.2.1) GO TO 920
50700 DH=DH-1.
50800 IF(DH.LT.10.) STOP
50900 GC TO 910
51000 920 HLT=0.
51100 A(2)=2.*3.14*RD*DH
51200 LC 1220 I=1,3
51300 H1(I)=A(I)*1.6*(TBDP-TO(I))/(1./PI(I))+B(I)/K(I)+1./PO(I))
51400 HLT=HLT+H1(I)
51500 1220 CCNTINUE
51600 THLD=NV*HLT
51700 HR=(TEMP-SLTENI)*1.8*(.54*TS*2.205+1.*WATR*2.205)
51800 WHL=.433072*1.05752** (TEMP)*(GASCH)*35.31
51900 TLR=THLD+HR+WHL
52000 TBTUC=TLR+NEE
52200 DPROF=DPROF+CH4LEN
52300 AECAP=AMCAP+ICCD
52400 GANDH=GANDH+DIGNC+MPC
52500 CAP=CAP+TCCC
52600 DH=DR*.3048
52700 RL=RD*.3048
52710 TBRH=CH4BTU
52730 HVAL=CH*BTU/GASCH
52740 H2O=GASCH/(1.-VP)-GASCH
52750 EKREQ=ENREQ+101UC
52800 WRITE(3,299) SLTEMP,TEMP,PH,RET,ANUSN,VED,TS,VS,ISR2H
52900 1299 FORMAT("DIGESTER PARAMETERS",/,
53000 10X,"INCOMING SLURRY TEMPERATURE...",7X,F5.2," DEG. C",/
53100 20X,"TEMPERATURE IN DIGESTER.....",7X,F5.2," DEG. C",/
53200 30X,"DIGESTER PH.....",7X,F5.2,/
53300 40X,"DIGESTER RETENTION TIME.....",7X,F5.2," DAYS",/
53400 50X,"UNDISSOCIATED AMMONIA.....",G12.5," MG/L NH3 AS N",/
53500 60X,"VOLATILE SOLIDS DESTROYED.....",G12.5," KG/HR",/
53600 70X,"TOTAL SOLIDS INPUT.....",G12.5," KG/HR",/
53700 80X,"VOLATILE SOLIDS INPUT.....",G12.5," KG/HR",/
53800 90X,"TOTAL SOLIDS IN EFFLUENT.....",G12.5," KG/HR")
53810 THLDJ=THLD*1055.
53820 HFIJ=HR*1055.
53830 WHIJ=WHL*1055.
53840 TBRJ=TLR*1055.
53900 WRITE(3,1300) SCOUT,NV,DH,ED,VD,TVD,THLDJ,HRJ,WHIJ,TBRJ
54000 1300 FORMAT(35X,"AT",F6.2" PERCENT SOLIDS",/,
54100 17X,14," DIGESTER(S) REQUIRED"/4X"WITH HEIGHT OF....."
54200 27X,F5.2," METERS"/4X,"AND RADIUS OF.....",7X,
54300 37X,F5.2," METERS"/4X,"VOLUME OF EACH DIGESTER.....",G12.5,
54400 4" CUBIC METERS"/4X,"TOTAL VOLUME OF DIGESTER(S)..."
54500 5G12.5," CUBIC METERS",/
54600 60X,"HEAT LOSS THROUGH DIGESTER(S).",G12.5," JOULES/HR",/
54700 70X,"HEAT REQUIRED TO HEAT SLURRY...",G12.5," JOULES/HR",/
54800 80X,"WATER VAPOR HEAT LOSS.....",G12.5," JOULES/HR",/
54900 90X,"TOTAL HEAT REQUIREMENT.....",G12.5," JOULES/HR")
54950 XINT=XINT*100.
54975 CH4JOU=CH4BTU*1055.
55000 WRITE(3,1301) ICCD,XINTP,TCCDA,NEC,DIGNC,GASCH,CO2GC,CH4CH,CH4JJJ

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55100 1301 FORMAT(4X,"DIGESTER CAPITAL COST.....",F12.2," $"/
55200 14X,"AMORTIZED AT "F5.2" PERCENT INTEREST...."F12.2" $/HR"/
55300 24X,"MIXING POWER COST....."F12.2" $/HR"/
55400 34X,"DIGESTER MAINTENANCE COST....."F12.2" $/HR"/
55500 44X,"GAS PRODUCTION"/
55600 54X,"TOTAL GAS....."G12.5" CUBIC METERS/HR"/
55700 64X,"CARBON DIOXIDE....."G12.5" CUBIC METERS/HR"/
55800 74X,"METHANE....."G12.5" CUBIC METERS/HR"/
55900 84X,"TOTAL METHANE ENERGY AVAILABLE"G12.5" JOULES/HR"/
55920 VS=VSREN
55970 TS=TSREN
56000 RETURN
56100 END
60000 SUBROUTINE CLEAN
60050 IMPLICIT REAL (R)
60100 CCMON4/UCON1,WATR,TS,VS,MOD,ASD,AMB,ALK,PH,DIGCON,VOLD,CH4CH,
60150 2 CO2CH,H2O,
60200 1 HVAL,CAP,ANCAP,ONDN,DENOP,ENREQ,ENRLE,WATNEC,WATIN,INCP
60400 CCMON4/UCON2/XINT,STC767,STC772,STC773,STC174,STC175,STC775,STC1,
60500 1 STC771,C4,C5,C6,C9,C10,SO1,AK1,RET,CONC,TEEP
60800 DATA MWHEA,SPHW,ALATH,GCLIP/61.08,0.5,1050.,15./
60850 DATA CMEA/0.50/
60900 C ****CALCULATE THE WEIGHT AND VOLUME OF MEA BEING CIRCULATED
61000 C ****WHEAC IS KG/HR SWEAC IS L/HR
61050 GASCH=CH4CH+CO2CH
61100 WEZAC=(CO2CH*35.31/359.)*1.5*MWHEA*.4536
61200 SWEAC=(WHEAC/0.2)
61300 C ****MAKEUP MEA IS 2% OF MEA CIRCULATED
61400 MEZEA=GASCH*1.1212E-04
61500 C ****COST OF MEA MAKEUP IS CMEA $/KG
61600 MEZEA=MEZEA*CMEA
61700 STEAM=1.1/3.785*SWEAC
61800 HVGCC=STEAM*(SPHW*(300.-212.)+(212.-120.)+ALATH)
61900 HZACC=1132000.*( (GASCH*24.*33.*35.31/(359.*2000.)) /150.)*.5
62000 1*STC1/STC773
62100 HEACCA=WEACCA*XINT/(1.-(1.+XINT)**(-GCLIP))/(360.*24)
62200 MEAONC=0.1*HEACCA/360./24.
62300 HECOST=HEACCA+MEAONC+MEZEA
62400 FERREQ=ENREQ+HVGCC
62500 ANCAP=ANCAP+HEACCA
62600 ONDN=ONDN+MEAONC+MEZEA
62700 CAP=CAP+HEACCA
62720 H2O=0.
62730 HVAL=35310.
62740 HVGCC=HVGCC*1055.
62800 WRITE(3,2920)GASCH,CH4CH,CO2CH,WHEAC,MWHEA,HVGCC,HZACC,HECOST,
62900 1HEACCA,MEAONC,MEZEA
63000 2920 FORMAT("1GAS CLEAN-UP SYSTEM OUTPUTS"/,
63100 14X,"TOTAL GAS IN....."G12.5," CUBIC METERS/HR"/
63200 24X,"METHANE....."G12.5," CUBIC METERS/HR"/
63300 34X,"CARBON DIOXIDE....."G12.5," CUBIC METERS/HR"/
63400 44X,"MEA REQUIRED....."G12.5," KG/HR"/
63500 54X,"MAKE-UP MEA....."G12.5," KG/HR"/
63600 64X,"HEAT REQUIRED....."G12.5," JOULES/HR"/
63700 74X,"CAPITAL INVESTMENT.....",F12.2," $"/
63800 84X,"HOURLY CCS1 TOTAL.....",F12.2," $"/
63900 94X," AMORTIZED CAPITAL.....",F12.2," $"/
64000 14X," OP. AND MAINT.....",F12.2," $"/
64100 24X," MAKE-UP MEA COST.....",F12.2," $"/
64150 CO2CH=0.
64200 RETURN

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64300      END
70000      SUBROUTINE CEN70
70020      COMMON/UCOH1/WATR,IS,VS,BIOD,ASH,ANN,ALK,PH,BIGCCG,VOLD,CHACH,
70040      1CC2CH,H2O,HVAL,CAP,AMCAP,OANDN,CPROF,ENREQ,ENREQ,ENREQ,WATR40,
70050      1 WATLN,INCF
70060      COMMON/UCOH2/XINT,STC767,STC772,STC773,STC774,STC775,STC776,
70080      1 STC777,C4,C5,C6,C9,C10,SO1,AK1,RET,CCRC,TEMP
70100      DATA C16,C17,QLIMT/10722.727,-214.545,51094.8/
70120      DATA UCCEXT,UCBLD,DARBA,EINST/125000.,40.,1000.,0.5/
70140      DATA BLIFE,DP,ANCH/10.,200.,.05/
70160      CCAKE=35.
70180      CCFN=.1
70200      CBN=XINT/(1.-(1.+XINT)**(-BLIFE))
70220      SFAFE=(C16+C17*CCAKE)/2.205
70240      NCRUN=INT(TS/SFAFE+.001)
70260      N=WATR/QLIMT
70280      IF(N.LT.1)N=1
70300      IF(N.LE.NCRUN) GO TO 400
70320      NCRUN=N
70340      400 NCACT=NCRUN+1
70360      SE=TS/NCRUN
70380      ALR=WATR/NCRUN
70400      C ****
70420      C **** CALCULATE CAPITAL COST
70440      C ****
70460      CCINT=NCACT*UCCEXT*STC1/STC174
70480      CCINST=EINST*CCINT
70500      CCBLD=DARBA*NCACT*UCBLD*STC1/STC174
70520      TCAPT=CCINT+CCINST+CCBLD
70540      CENCCA=TCAPT*CR/(360.*24.)
70560      C ****
70580      C **** POWER REQUIREMENTS AND COSTS
70600      C ****
70620      P=UP*NCRUN*C10
70640      EC=HP*NCRUN*C5*CE
70660      C **** MAINTENANCE COST
70680      CENMC=(ANCH*TCAPT)/(360.*24.)
70700      HCCOST=CENCCA+P+CE+CE+CE
70720      CCEN=CCEN/100.
70740      CCAKE=CCAKE/100.
70760      C ****
70780      C **** CALCULATE OUTPUTS FROM CENTRIFUGES
70800      C ****
70820      CENTS=CCEN*(WATR-IS*(1.-CCAKE)/CCAKE)/(1.-CCEN-
70840      1 CCEN*(1.-CCAKE)/CCAKE)
70860      CENMO=CENTS*(1.-CCEN)/CCEN
70880      CEN=CENTS+CENMO
70900      CAKETS=TS-CENTS
70920      CAKEVS=CAKETS*(VS/TS)
70940      CAKCHO=CAKETS*(1.-CCAKE)/CCAKE
70960      CAKE=CAKETS+CAKCHO
70980      ENREQ=ENREQ+P
71000      AMCAP=AMCAP+CENCCA
71020      CANDN=OANDN+EC+CENMC
71040      CAP=CAP+TCAPT
71060      BIOD=BIOD*CENTS/TS
71080      ASH=ASH*CENTS/TS
71100      ANN=ANN*CENTS/TS
71110      VS=VS*CENTS/TS
71120      TS=CENTS
71140      WRITE(3,70) NCRUN,NCACT,SH,ALR

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71160 70 FORMAT(//," CENTRIFUGE OUTPUTS",/
71180 14X,"NO. CENTRIFUGES RUNNING.....",8X,14," UNITS",/
71200 14X,"NO. CENTRIFUGES INCLD. BACKUP.",8X,14," UNITS",/
71220 14X,"SOLIDS FLOW RATE IN EACH.....",F12.2," KG/H2",/
71240 14X,"LIQUID FLOW RATE IN EACH.....",F12.2," L/HR")
71260 WRITE(3,7*) TCAPT,HCCST,CENCCA,PC,CENNC
71280 71 FORMAT(//," COST SUMMARY FOR CENTRIFUGES",/
71300 14X,"CAPITAL COST TOTAL.....",F12.2," $",/
71320 14X,"HOURLY COST TOTAL.....",F12.2," $/HR",/
71340 14X," AMORTIZED CAPITAL.....",F12.2," $/HR",/
71360 14X," POWER COST.....",F12.2," $/HR",/
71380 14X," MAINTENANCE COST.....",F12.2," $/HR",/)
71400 RETURN
71500 END
80000 SUBROUTINE FINAL
80100 X=1.
80200 RETURN
80300 END
85000 SUBROUTINE HEA1(THR,IGC)
85100 CCMON/UCCH1/WATR,TS,VS,BIGD,ASH,ANN,ALK,PH,DIGCON,VOLD,CH4CH,
85200 1 CO2CH,H2O,HVAL,CAP,AMCAP,OANDN,DFROP,LBRQ,ZN2LD,WATN2D,
85250 1 WATN,INCP
85300 CCMON/UCCH2/XINT,STC767,STC772,STC773,STC774,STC775,STC776,
85400 1 STC771,C4,C5,C6,C9,C10,SO1,AK1,BLT,CCNC,TEMP
85500 GASCH=CH4CH+CO2CH
85600 FRAC=CH4CH/GASCH
85650 H2OPR=H2O/GASCH
85700 GASCH=GASCH-THR/HVAL
85800 PRINT(6,*)"INCLUDE GAS CLEANUP? 1=YES 0=NO",
85900 READ(5,*) IGC
86000 IF(IGC.NE.1) GO TO 100
86100 GASCH=GASCH/(1.+7044.46*(1.-FRAC)/HVAL)
86200 100 BTUUSE=ENRFX-GASCH*HVAL
86300 ENREN=GASCH*HVAL
86400 ENREQ=ENREQ-BTUUSE
86500 CH4CH=GASCH*FRAC
86600 CO2CH=GASCH-CH4CH
86650 H2Q=H2OPR*GASCH
86700 STEAM=BTUUSE/1186.
86800 CCBOI=10.** (ALOG10(STEAM)*.0648+1.75)*STC1/STC772
86900 CCBOIA=CCBOI/360./24.*XINT/(1.- (1.+XINT)** (-20.))
87000 BOIOP=.03*CCBOI/360./24.
87100 WAT=STEAM/60./8.34
87200 CCWAT=10.** (ALOG10(WAT)*1.42+2.239)*STC1/STC772
87300 CCWATA=CCWAT/24./360.*XINT/(1.- (1.+XINT)** (-20.))
87400 WATOP=.02*CCWAT/24./360.
87500 CAP=CAP+CCBOI+CCWAT
87600 AMCAP=AMCAP+CCBOIA+CCWATA
87700 OANDN=OANDN+WATOP+BOIOP
87800 WAT=WAT*60.*3.785
87850 DFROP=DFROP-BTUUSE*C9/1.806
87900 AJUSE=BTUUSE*1055.
87950 WRITE(3,27) AJUSE,CCBOI,CCBOIA,BOIOP,WAT,CCWAT,CCWATA,WATOP
88000 27 FORMAT("1BOILER SYSTEM OUTPUTS",/
88100 14X,"BOILER CAPACITY.....",G12.5," JOULES/H2",/
88200 24X," CAPITAL COST.....",F12.2," $",/
88300 34X," AMORTIZED CAPITAL.....",F12.2," $/HR",/
88400 44X," OP. AND MAINT.....",F12.2," $/HR",/
88500 54X,"DENMINERALIZATION SYSTEM",/
88600 64X,"CAPACITY.....",G12.5," L/HR",/
88700 74X," CAPITAL COST.....",F12.2," $"/

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88800      84X," AMORTIZED CAPITAL.....",F12.2," $/HR"/
88900      94X," OP. AND MAINT.....",F12.2," $/HR"/)
89000      RETURN
89001      END
89100      SUBROUTINE CAUST
89200      COMMON /UCOM1/MATR,IS,VS,DIOD,ASH,A1H,ALX,PH,DIGCUB,VOLD,CHUC1,
89300      1 CO2CH,H2O,HVAL,CAP,ANCAP,UAHDA,UPROF,LENEL,ENH4,ATMED,
89400      2 MATIN,INCF
89500      COMMON/UCOM2/KINT,SIC767,STC772,STC773,STC174,STC175,STC775,SAC1,
89600      1 STC771,L4,C5,C6,C9,L10,S01,AK1,BLF,CCNC,TAMP
89700      CUB(X,BLF)=X/(1.-(1.+X)**(-BLF))
89800      TVOL=IS*.064*2.2
89900      NTNK=INT(TVOL*.4./COJ0.+1.)
90000      TNKCC=(150000.*NTNK)*.TC1/STC775
90100      TNKCCA=TNKCC*LEN(LKINT,10.)/360./24.
90200      CHEMF=TS*.03*2.2
90300      CAUS=15%.*CHEMF/20)0.
90400      CAUMMC="NAKCA".2
90500      CAYOM=O1NUM+CAUMMC+CAUS
90600      CAP=CAP+TNKCC
90700      ANCAP=ANCAP+TNKCCA
90800      CLLEF=CHEMF/2.2
90900      WRITE(3,71)NINX,"NAKCC,TNKCCA,CAUMMC,CHEMF,CAUS
91000      71 FORMAT(*,"CAUSTIC TREATMENT PARAMETERS",//
91100      1" NUMBER OF TANKS REQUIRED....."10X,12,/
91200      2" CAPITAL COST....."F12.2," $"/,
91300      3" AMORTIZED AT 10. PERCENT....."F12.2," $/HR"/,
91400      4" OP. AND MAINT.....",F12.2," $/HR"/,
91500      5" CAUSTIC FLOW RATE.....",F12.0," KG/HR"/,
91600      6" CAUSTIC COSI....."F12.2," $/HR"/)
91700      RETURN
91800      END

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4. Sample Output

RESIDUAL DIGESTION SIMULATION

PLANT CAPACITY.....	1000 TONS/DAY
DIGESTER INPUT CONCENTRATION.....	12.90 PERCENT
RETENTION TIME REQUESTED.....	7.00 DAYS
RECYCLE RATE FOR CENTRATE.....	0.00 PERCENT

PARAMETERS USED FOR SOLIDS DESTRUCTION

K.....	.25 PER DAY
COMPOSITION OF FEED MATERIAL	
WATER.....	15.00 PERCENT
TOTAL SOLIDS.....	85.00 PERCENT
VOLATILE SOLIDS.....	90.00 PERCENT T.S.
BIODEGRADABLE SOLIDS.....	77.90 PERCENT V. S.
CAUSTIC TREATMENT PARAMETERS	
NUMBER OF TANKS REQUIRED.....	10
CAPITAL COST.....	1673844.95 \$
AMORTIZED AT 10. PERCENT.....	31.53 \$/HR
OP. AND MAINT.....	6.31 \$/HR
CAUSTIC FEE RATE.....	964. KG/HR
CAUSTIC COST.....	167.49 \$/HR

MATERIAL BALANCE AFTER DATA INITIALIZATION

WATER IN FEED STREAM.....	5669.0	L/HR
TOTAL SOLIDS IN FEED STREAM.....	32124.	KG/HR
VOLATILE SOLIDS IN FEED STREAM.....	28912.	KG/HR
BIODEGRADABLE SOLIDS IN FEED STREAM	22262.	KG/HR
ASH IN FEED STREAM.....	3212.4	KG/HR
AMMONIA NITROGEN IN FEED STREAM....	0.	KG/HR AS N

ECONOMIC SUMMARY

TOTAL CAPITAL REQUIRED.....	1673844.95 \$
AMORTIZED AT 10.00 PERCENT.....	31.53 \$/HR
OPERATION AND MAINTENANCE.....	173.80 \$/HR
VALUE OF RECOVERED MATERIAL AND GAS	0.00 \$/HR

ENERGY BALANCE

ENERGY REQUIRED FOR PROCESSING.....	0.	JOULES/HR
ENERGY AVAILABLE IN SYSTEM.....	.57840E+12	JOULES/HR

MIX TANK OUTPUTS

TOTAL WATER REQUIREMENTS FROM RECYCLE AND
MAKEUP FOR 12.90 PERCENT INPUT SOLIDS:

.21123E+06 L/HR
55881. GAL/HR

NUTRIENT COST..... 21.47 \$

MATERIAL BALANCE AFTER SLURRY MIXING

WATER IN FEED STREAM.....	.21690E+06	L/HR
TOTAL SOLIDS IN FEED STREAM.....	32124.	KG/HR
VOLATILE SOLIDS IN FEED STREAM.....	28912.	KG/HR
BIODEGRADABLE SOLIDS IN FEED STREAM	22262.	KG/HR
ASH IN FEED STREAM.....	3212.4	KG/HR
AMMONIA NITROGEN IN FEED STREAM....	1.4172	KG/HR AS N

ECONOMIC SUMMARY

TOTAL CAPITAL REQUIRED.....	1770159.97	\$
AMORTIZED AT 10.00 PERCENT.....	32.84	\$/HR
OPERATION AND MAINTENANCE.....	195.54	\$/HR
VALUE OF RECOVERED MATERIAL AND GAS	0.00	\$/HR

ENERGY BALANCE

ENERGY REQUIRED FOR PROCESSES.....	.36346E+09	JOULES/HR
ENERGY AVAILABLE IN SYSTEM.....	.57840E+12	JOULES/HR

DIGESTER PARAMETERS

INCOMING SLURRY TEMPERATURE...	10.00	DEG. C
TEMPERATURE IN DIGESTER.....	58.00	DEG. C
DIGESTER PH.....	6.80	
DIGESTER RETENTION TIME.....	7.00	DAYS
UNDISSOCIATED AMMONIA.....	.21212	MG/L NH3 AS N
VOLATILE SOLIDS DESTROYED.....	13095.	KG/HR
TOTAL SOLIDS INPUT.....	32124.	KG/HR
VOLATILE SOLIDS INPUT.....	28912.	KG/HR
TOTAL SOLIDS IN EFFLUENT.....	19029.	KG/HR

AT 8.07 PERCENT SOLIDS

5 DIGESTER(S) REQUIRED		
WITH HEIGHT OF.....	14.63	METERS
AND RADIUS OF.....	13.49	METERS
VOLUME OF EACH DIGESTER.....	8367.2	CUBIC METERS
TOTAL VOLUME OF DIGESTER(S)...	41836.	CUBIC METERS
HEAT LOSS THROUGH DIGESTER(S)...	.39309E+10	JOULES/HR
HEAT REQUIRED TO HEAT SLURRY...	.47081E+11	JOULES/HR
WATER VAPOR HEAT LOSS.....	.44331E+10	JOULES/HR
TOTAL HEAT REQUIREMENT.....	.55445E+11	JOULES/HR
DIGESTER CAPITAL COST.....	4764291.93	\$
AMORTIZED AT 10.00 PERCENT INTEREST....	64.77	\$/HR
MIXING POWER COST.....	31.45	\$/HR
DIGESTER MAINTENANCE COST.....	5.51	\$/HR

GAS PRODUCTION		
TOTAL GAS.....	10702.	CUBIC METERS/HR
CARBON DIOXIDE.....	5269.4	CUBIC METERS/HR
METHANE.....	5432.4	CUBIC METERS/HR
TOTAL METHANE ENERGY AVAILABLE	.20237E+12	JOULES/HR

MATERIAL BALANCE AFTER DIGESTION

WATER IN FEED STREAM.....	.21690E+06	L/HR
TOTAL SOLIDS IN FEED STREAM.....	19029.	KG/HR
VOLATILE SOLIDS IN FEED STREAM.....	15817.	KG/HR
BIODEGRADABLE SOLIDS IN FEED STREAM	9166.9	KG/HR
ASH IN FEED STREAM.....	3212.4	KG/HR
AMMONIA NITROGEN IN FEED STREAM.....	1.4172	KG/HR AS N

DIGESTER EQUILIBRIUM CONDITIONS

ALKALINITY ADDED WITH INPUT STREAMS	.30717E-03	G/L FEB DAY
DIGESTER PH.....	6.80	
EFFLUENT SLURRY CONCENTRATION.....	8.07	PERCENT
TOTAL DIGESTER VOLUME.....	41836.	CUBIC METERS

GAS STREAM CONSTITUENTS

METHANE.....	5432.4	SCM/HR
CARBON DIOXIDE.....	5209.4	SCM/HR
WATER VAPOR.....	2333.3	SCM/HR
HEAT VALUE OF GAS.....	.18910E+06	JOULES/SCM

ECONOMIC SUMMARY

TOTAL CAPITAL REQUIRED.....	6534451.90	\$
AMORTIZED AT 10.00 PERCENT.....	97.01	\$/HR
OPERATION AND MAINTENANCE.....	232.50	\$/HR
VALUE OF RECOVERED MATERIAL AND GAS	383.63	\$/HR

ENERGY BALANCE

ENERGY REQUIRED FOR PROCESSES.....	.59604E+11	JOULES/HR
ENERGY AVAILABLE IN SYSTEM.....	.20237E+12	JOULES/HR

BOILER SYSTEM OUTPUTS

BOILER CAPACITY..... .55445E+11 JOULES/HR

CAPITAL COST..... 115719.71 \$

AMORTIZED CAPITAL..... 1.57 \$/HR

OP. AND MAINT..... .40 \$/HR

DEMINEERALIZATION SYSTEM

CAPACITY..... 20111. L/HR

CAPITAL COST..... 169213.19 \$

AMORTIZED CAPITAL..... 2.70 \$/HR

OP. AND MAINT..... .39 \$/HR

MATERIAL BALANCE AFTER BOILER

WATER IN FEED STREAM.....	.21690E+06	L/HR
TOTAL SOLIDS IN FEED STREAM.....	19029.	KG/HR
VOLATILE SOLIDS IN FEED STREAM.....	15817.	KG/HR
BIODEGRADABLE SOLIDS IN FEED STREAM	9166.9	KG/HR
ASH IN FEED STREAM.....	3212.4	KG/HR
AMMONIA NITROGEN IN FEED STREAM....	1.4172	KG/HR AS N

GAS STREAM CONSTITUENTS

METHANE.....	3944.0	SCM/HR
CARBON DIOXIDE.....	3325.6	SCM/HR
WATER VAPOR.....	1694.0	SCM/HR
HEAT VALUE OF GAS.....	.18910E+06	JOULES/SCM

ECONOMIC SUMMARY

TOTAL CAPITAL REQUIRED.....	6819384.79 \$
AMORTIZED AT 10.00 PERCENT.....	101.48 \$/HR
OPERATION AND MAINTENANCE.....	233.30 \$/HR
VALUE OF RECOVERED MATERIAL AND GAS	278.52 \$/HR

ENERGY BALANCE

ENERGY REQUIRED FOR PROCESSES.....	.41566E+10 JOULES/HR
ENERGY AVAILABLE IN SYSTEM.....	.14692E+12 JOULES/HR