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LANDFILL GAS CONVERSION TO LNG and LCO_2

PHASE II
Contract No 725089 (CH10982)
PHASE II FINAL REPORT

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prepared for

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EXECUTIVE SUMMARY

This report summarizes work on the development of a process to produce LNG (liquefied methane) for heavy vehicle use from landfill gas (LFG) using Acrion's CO_2 wash process for contaminant removal and CO_2 recovery. Work was done in the following areas: 1) production of natural gas pipeline methane for liquefaction at an existing LNG facility, 2) production of LNG from sewage digester gas, 3) the use of mixed refrigerants for process cooling in the production of LNG, liquid CO_2 and pipeline methane, 4) cost estimates for an LNG production facility at the Arden Landfill in Washington PA.

Process designs and economics were developed to produce pipeline gas and liquid carbon dioxide (CO_2) from landfill gas (LFG) using the Acrion CO_2 wash process. The patented Acrion CO_2 wash process uses liquid CO_2 to absorb contaminants from the LFG. The process steps are compression, drying, CO_2 wash contaminant removal and CO_2 recovery, physical solvent residual CO_2 removal to pipeline specifications. Installed capital cost is $5.3 million, annual operating costs $1.0 million, and project payback is 3.1 years with methane at $2.00 per MMBtu and liquid CO_2 at $40/ton. This design is the basis for determining the economic feasibility of liquefying natural gas at large LNG peak shaving facilities, swapping lower cost landfill methane injected into the distribution system near the landfill for natural gas removed from the supply pipe at the peak shaving plant.

Pipeline natural gas specifications have been compiled and are reported. Two pipeline gas characteristics crucial to conversion of landfill gas to pipeline methane are heating value (Btu/SCF) and total inerts. These characteristics are not independent and are the most difficult to achieve in the upgrade of landfill methane to pipeline specifications. Heating value can be increased by propane injection, and may prove economic if the amount required is less than about 1 to 2 vol%. Injection of processed landfill methane into distribution and transmission pipelines must be examined carefully on a case-by-case basis. and lengthy negotiations with the pipeline owner can be anticipated.

Vandor + Vandor, an independent subcontractor, studied "Wheeling" landfill methane for LNG. Vandor's work examined the feasibility of utilizing excess liquefaction capacity at large peak shaving plants to produce LNG, and distribution of LNG from the peak shaving plant(s) with Maryland. The concept appears feasible.

Sewage digester gas as a methane source for production of liquid methane has been investigated. AplusB, Inc., an independent subcontractor, studied sewage digester gas and reviewed commercial or near-commercial bio-digester technologies. Process designs were developed to convert digester gas to LNG at two levels of H_2S in the raw gas, 100 ppm and 600 ppm. Economics appear favorable for a 21,500 gal/day production facility with simple paybacks of 3 and 3.3 years. However Acrion does not intend to actively pursue sewage digester gas as a methane source for LNG heavy duty truck fuel for several reasons: 1) digester gas is tightly integrated into the energy balance of most waste water treatment plants and conversion of digester methane to LNG for offsite use would require procurement of an alternative less expensive energy source to replace it; and 2) high H_2S levels at some wastewater treatment plants increase costs and 3) there is much greater opportunity to develop LNG projects at municipal landfills.

Mixed refrigerant systems have been compared with a conventional cascade refrigeration system for the LNG and liquid CO_2 designs developed in Phase I and the pipeline gas and liquid CO_2 design developed in Phase II. For the low temperature LNG and liquid CO_2 design the power requirement for the mixed refrigerant and cascade refrigeration systems were the same but the capital cost of the mixed refrigerant system was about 5% lower due to the presence of a single compression unit. For the warmer refrigeration requirements of the pipeline gas and CO_2 refrigeration system (-70°F), the mixed refrigerant system had a 19% lower power requirement than the cascade system and a 9% lower capital cost.

A cost estimate for gas compression and cleanup to supply a small LNG production facility to fuel refuse vehicles at the Arden Landfill in Washington PA was developed. Vendor quotes for a 1200 gal/day facility were acquired. The estimated plant cost is $860,000.

The next step in the development process is to define the size and scope of a project at the Arden landfill. Acrion would pursue the task of firming the cost estimate and identifying lower cost technologies and sources for trace CO_2 removal. We would also identify the market for co-produced CO_2 in the area. Acrion's pilot LFG to liquid CO_2 pilot unit, currently under construction for placement at the NJ Ecocomplex, can supply high pressure contaminant free methane enriched gas which can be further processed to approximately 800 gal/day of LNG.

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A. INTRODUCTION

This report summarizes work completed during the Phase II period of January 1999 through April 2000 under contract 725089 entitled "Landfill Gas Conversion to LNG and Liquid CO_2." This Phase II research program was proposed and awarded under Brookhaven National laboratory RDP #723418 *"Liquefied Natural Gas as a Heavy Vehicle Fuel."*

The objective is to examine the feasibility of using municipal landfill gas and other biogases as feedstock for the production of liquefied methane and liquid carbon dioxide. Municipal landfill gas (LFG) is a mixture of methane and carbon dioxide, roughly 50:50, and up to several thousand parts per million of troublesome contaminants. Reliable, economic removal of contaminants from LFG is a barrier to widespread utilization of LFG methane and carbon dioxide in most potential end markets.

Acrion has developed and patented technology [US Patents 5,681,360 and 5,842,357, Landfill Gas Recovery, 28 Oct 97 and 1 Dec 98] which removes contaminants from LFG and produces clean mixtures of methane and carbon dioxide suitable for processing to a variety of finished products. The raw landfill gas is compressed, dried, and cooled to condense a majority of the CO_2. A portion of the liquid CO_2 condensate is used to absorb landfill gas contaminants; the remainder is a commercial liquid CO_2 product. Acrion's contaminant removal and CO_2 recovery technology integrated with conventional CO_2 separation, solvents and membranes, can produce a methane product for natural gas pipelines and a commercial liquid CO_2 product. The methane can be further processed to remove trace amounts of CO_2 and produce liquefied natural gas (LNG). One option studied in this work is to have the LNG production step done a an existing LNG production facility by producing pipeline methane which could be traded for pipeline gas at the existing production facility. A typical landfill gas feed is characterized below in Table A.1. Sewage digester gas has a higher methane content and lower CO_2 content than LFG and its major contaminant is typically hydrogen sulfide (H_2S).

Table A.1

Landfill Gas Feed Properties

Temperature, °F	70
Pressure, psia	14.7
Methane	50-55%
Carbon Dioxide	40-45%
Nitrogen	0-5%
Contaminants	.001-.1%
Water	saturated

This report is organized as follows. Section B reports progress on producing pipeline gas for use at an existing LNG facility. Section C reports on the use of digester gas as a source of methane for LNG production. Section D reports on the use of mixed refrigerants for process cooling and Section E reports on small LFG cleanup economics.

B. Subtask 2.1
Pipeline Gas and Liquid CO2 Process Flowsheet Development

B.1 Pipeline Methane from Landfill Gas

Acrion has completed process designs to produce pipeline gas and carbon dioxide from landfill gas using Acrion's in-situ carbon dioxide wash technology. The design is part of Task 2 objective to determine the economic feasibility of liquefying natural gas at large LNG peak shaving facilities, swapping low(er) cost landfill methane injected into the distribution system near the landfill for natural gas removed from the supply pipe at the peak shaving plant. This section a) reviews the process design an economic analysis for production of pipeline methane and commercial liquid carbon dioxide, b) presents the results of inquires to natural gas transporters and distributors regarding gas quality requirements for pipeline injection, c) summarizes a preliminary study of swapping pipeline quality methane at a landfill for liquid natural gas produced at a large scale peak shaving plant (swapping gaseous Btu's for liquid Btu's), and d) summarizes a preliminary study of swapping electricity produced with landfill gas for liquid natural gas produced at a large scale peak shaving plant (swapping kWh's for liquid Btu's).

B.1.1 Pipeline Gas Production from LFG

The design basis is given in Table B.1 and product characterization in Table B.2. Methane recovery is 99.95%. Carbon dioxide recovery is 84.6%. Other effluent streams generated by the process include condensate water and 0.227 MMSCFD (24.9 lb-moles/hr) of contaminant rich carbon dioxide gas containing 0.5% methane. This contaminant rich carbon dioxide is incinerated.

Table B.1
Design Basis
Pipeline Methane and Liquid CO_2 from Landfill Gas

LFG flow	4 MMSCFD* = 439 lb-mole/hr (dry)
LFG pressure	14.7 psia (ambient)
LFG composition (dry basis)	54% methane (CH_4) 45% carbon dioxide (CO_2) 1% nitrogen (N_2) trace contaminants water dew point 70°F

*(million standard cubic feet per day)

Table B.2

Product Characterization

	Pipeline Gas	Liquid CO2
Flow	2.25 MMSCFD	88.3 tons/day
	247 lb-mol/hr	167.2 lb-mol/hr
Pressure	485 psia	300 psia
Temperature		-1.4°F
Composition		
CH4	96.0%	1 ppm
CO2	2.2%	99.99%
N2	1.8%	
Btu/SCF	960	
State	gas	liquid

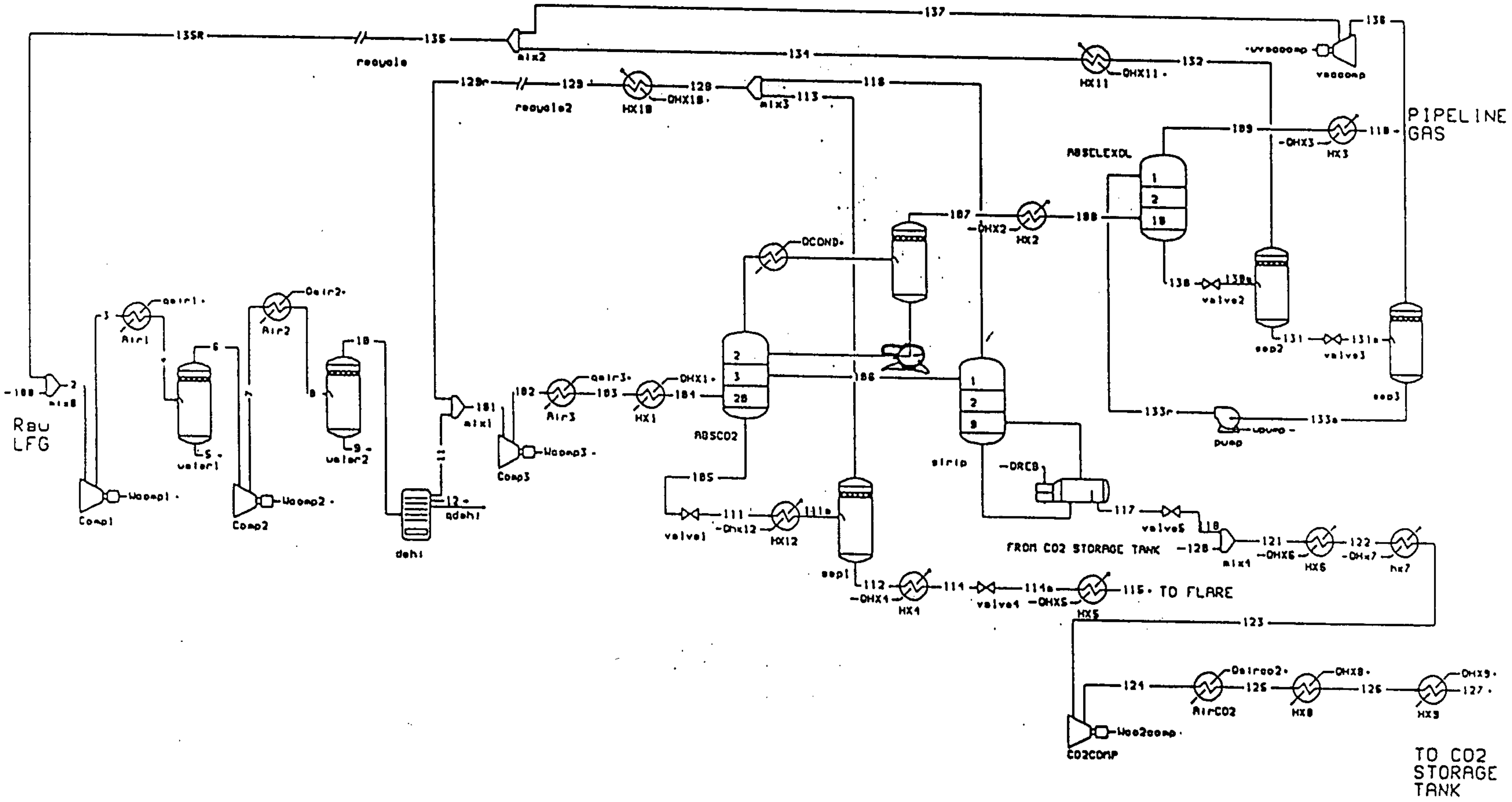
B.1.2 Process Description

The process flowsheet is shown in Figure B.1. The detailed flowsheets and a material and energy balance are in Appendix A. The raw LFG is compressed to 500 psig in three stages of compression. Air intercoolers and knock out drums remove water after each stage of compression. The compressed gas passes through a Sulfatreat bed to remove H2S and a mol sieve drying bed to remove trace water. The compressed and dried gas (stream 103 in Figure B.1) is cooled by heat exchange with pipeline gas product, recycle gas and flare gas in HX1. The gas enters the absorber (ABSCO2) where the contaminants are washed out with liquid carbon dioxide. A portion of the carbon dioxide in the overhead vapor from the absorber is condensed. The condensed carbon dioxide is pumped (pump not shown on flowsheet) back to the top of the absorber tower to be used as absorbent. Part of the carbon dioxide absorbent is removed near the top of the absorber. This stream (stream 106), free of contaminants, is sent to the methane stripper to remove dissolved methane. The rest of the carbon dioxide absorbent passes through the absorber and leaves the bottom of the column enriched in contaminants.

This contaminant enriched stream (stream 105) is flashed down to 180 psia and warmed by heat exchange with the condenser to vaporize the dissolved methane in the liquid. The vapor (stream 113) is recycled back to the last stage of landfill gas compression. The remaining liquid (stream 112) is vaporized, warmed, expanded to near atmospheric pressure and flared. This gas does not have sufficient methane to be flammable, therefore a small amount of raw landfill gas is mixed with it to ensure combustion.

Figure B.1

LFG to Pipeline Gas and Carbon Dioxide



The methane stripper operates at 180 psia and removes methane from the contaminant free carbon dioxide down to the 1 ppm level. The vapor (stream 116) from the stripper is recycled back to the last stage of landfill gas compression. The liquid (stream 117) from the bottom of the stripper is the pure carbon dioxide product.

Vapor from the top of the contaminant absorber(stream 107), containing approximately 25% carbon dioxide is warmed to 0°F by heat exchange with the feed and is sent to the SELEXOL column where the remaining carbon dioxide is removed. Gas (stream 109) leaves the top of the SELEXOL column with 4% nitrogen and carbon dioxide. It is warmed to near ambient temperature and is taken as the pipeline gas product. The SELEXOL solvent at the bottom of the column (stream 130) is flashed to near atmospheric pressure, and vacuum flashed to regenerate the solvent. It is then pumped back to the top of the SELEXOL column. Vapor from the two flashes is recycled back to the landfill gas feed.

Liquid carbon dioxide product is used as a refrigeration source for the contaminant absorber condenser. It is flashed down to 78 psia where it provides cooling by boiling at -67°F in heat exchanger HX6. Vapor from the heat exchanger is warmed in HX7, compressed to 300 psi, and cooled to near ambient temperature. Gas would pass through a carbon bed at this point if it were needed. The gas is cooled in HX8 (the other side of HX7), condensed with refrigeration in HX9 and is stored in a liquid carbon dioxide storage tank at 300 psia and 0°F.

B.1.3 Process Equipment Sizing and Economics

Specifications and cost of major pieces of capital equipment are shown in Table B.3.. Capital and operating costs and estimated revenues are shown in Table B.4.

B.2 Pipeline Company Gas Requirements

Results of inquiry to several local gas distribution companies and gas transmission companies are presented in Table B.5. Most pertinent to this investigation is Baltimore Gas & Electric, which has excess capacity for methane liquefaction in the greater Baltimore, Maryland area. A recent survey of pipeline gas quality specifications was conducted by Gas Research Institute [1]. At least 26 organizations were responsive to each of the various gas quality criteria. Results of GRI's survey which impact landfill gas processing to pipeline specification(s) are shown in Table B.6; probably the most important parameter is Btu content, or heating value. Since no distinction is made in the reported results pertaining to "wet" gas or "dry" gas, a safe assumption is that virtually all pipelines require a minimum Btu content of about 967 to 970 Btu/SCF (dry basis).
Two pipeline gas characteristics are highlighted in Table B.5: 1) heating value, and 2) total inerts. These characteristics are not independent, and are the most difficult to achieve in the upgrade of landfill methane to pipeline specifications. Consider heating value. Pipeline gas with 96% methane, 4% inerts, has a heating value of 972 Btu (all heating values cited herein are higher heating values (water as liquid), and refer to one standard cubic foot of gas, SCF), just barely above 967 Btu, and more than a tad below 1000 Btu. To achieve 1000 Btu, this gas would need to contain at least 3.7% ethane (4% inert, 3.7% ethane, 92.3% methane), or it would need to contain at least 1.9% propane

Table B.3
Equipment Cost
4 MMSCFD Raw LFG
Pipeline Gas + Liquid Carbon Dioxide from LFG
Acrion Contaminant Removal Process
and Selexol Final CO2 Removal

Tag	Service	Quantity	Size		Mat. + Press.	Equipment Cost, $K	Installed Cost 1995 $K
Compressors							
Comp1	Landfill Gas Stage 1	1	473	hp		$243	$522
Comp2	Landfill Gas Stage 2	1	486	hp		$247	$530
Comp3	Landfill Gas Stage 3	1	486	hp		$247	$530
CO2Comp	CO2 Compressor	1	210	hp		$150	$324
vaccomp	Vaccuum Pump	2	5	hp		$38	$82
						$925	$1,988
Refrigerators							
HX9	Product CO2 Condenser (-20F)	1	94.4	Tons		$309	$472
							$472
Exchangers							
HX2	Feed Cooler	1	79	ft^2	SS	$13	$40
HX3+HX11	Feed Cooler	1	342	ft^2	CS	$13	$42
HX4	Contam. CO2 evap.	1	55	ft^2	SS	$10	$31
HX5+HX10	Feed Cooler	1	288	ft^2	SS	$31	$99
HX6	CO2 Condenser	1	548	ft^2	Al	$41	$130
HX7/8	CO2 cooler/heater	1	155	ft^2	Al	$17	$54
HX12	Contam. CO2 evap/cond	1	338	ft^2	SS	$35	$111
Reboiler	Stripper Reboiler	1	143	ft^2	Al	$16	$51
							$558
Pumps							
	Liquid CO2 Absorbent	2	0.5	hp	SS	$10	$33
pump	Selexol Solvent	2	42	hp	CS	$21	$68
							$101
Towers			Dia	Ht			
ABSCO2	Contaminent Absorber	1	2.'	35'	CS 500 psi	$39	$162
strip	Methane Stripper	1	1'	20'	KCS 200 psi	$10	$40
ABSELEXOL	CO2 Absorber	1	2'	40'	CS 500 psi	$40	$168
							$371
Vessels			Dia	Ht			
	CO2 Reflux	1	2.0'	9'	SS 500 psi	$43	$179
separator1	CO2 Flash	1	1.5'	5'	KCS 200 psi	$5	$20
separator2	Selexol Flash	1	3.0'	14'	CS 20 psi	$9	$39
separator3	Selexol Vacuum Flash	1	3.0'	14'	CS 7 psi	$9	$39
							$277
						Subtotal	$3,766
			Dia	Ht			
Dehydration	Mol Sieve Beds+ Heater	2	2.5	13'	CS 510psi	$63	$191
H2S Removal	Sulfatreat Beds	2	5'	10'	CS 510psi	$61	$254
Storage	Liquid CO2 Storage Tank	1	60	KGal	CS 300psi	$134	$280
						Total	$4,491

CS= Carbon Steel
SS= Stainless Steel
KCS = Killed Carbon Steel

Table B.4

Capital, Operating Costs, and Revenues
Landfill Gas Recovery for Pipeline Gas and CO2

RAW LFG CONSUMPTION		4 MMSCFD	54% CH4
OPERATING PARAMETERS			
Methane Recovery		99.95%	
Carbon Dioxide Recovery		84.60%	
Pipeline Gas Pressure		485 psia	
Methane Content of Pipeline Gas		96.0%	
Onstream Factor, 350 days/year		96%	
PRODUCT	PRICE		
Pipeline Gas	$2.00 /MMBtu		
Liquid CO2	$40/Ton		
Power Required	1,454 kW		
CAPITAL COSTS		Installed	
Equipment		Cost	
Compressors		1,988,000	
Refrigerators		472,000	
Columns		371,000	
Vessels		277,000	
Dehydration		191,000	
Heat Exchangers		558,000	
H2S Removal		254,000	
Pumps		101,000	
CO2 Storage Tank		280,000	
Capital Cost		4,492,000	
Contingency @ 18%		809,000	
Total Capital Costs		5,301,000	
ANNUAL OPERATING EXPENSES			
Electric Power	4.5 ¢/kWh	549,612	
SulfaTreat		10,933	
Labor ($12/hr, 2 operators/day)		67,000	
Labor Overhead (100% of labor)		67,000	
Maintenance Materials (2% of capital)		106,000	
Maintenance Labor (3% of capital)		159,000	
Taxes & Insurance (1.5% of capital)		80,000	
Total Operating Costs		1,040,000	
ANNUAL INCOME	Amount	Daily	Annual
Pipeline Gas	2,250 MSCFD	4,318	1,511,000
Liquid CO_2	88 TPD	3,533	1,237,000
Total Income		7,851	2,748,000
PAYBACK PERIOD	3.1 Years	ACRION TECHNOLOGIES August 18, 1999	

(4% inert, 1.9% propane, 94.1% methane), or some acceptable combination of ethane, propane and/or higher hydrocarbon.

In contrast with natural gas, landfill gas contains no significant ethane, propane or higher hydrocarbons, generally far less than 1% total; thus heating value of product gas is totally dependent on methane. Furthermore, since landfill gas is 50% methane, 50% CO2, carbon dioxide removal concentrates inerts such as nitrogen and oxygen with the methane product. Thus, 2% inerts in raw landfill gas becomes (roughly) 4% inerts in product methane, not including any residual carbon dioxide. Only slight contamination of raw landfill gas with inerts (air) can prevent attainment of pipeline heating value requirement. Figure B.2 plots CO2 content of product methane needed to attain 967 Btu/SCF heating value versus the inert content of raw landfill gas feedstock containing 50% methane. (None of the inert gas is assumed to be separated from methane during removal of contaminants and CO2 from landfill gas.) For example, if raw landfill gas contains 2% inerts, CO2 separation must achieve less than 1% CO2 in the product gas to produce 967 Btu/SCF product. If raw landfill gas contains 2.5% inerts, 967 Btu/SCF product is impossible to achieve. This conclusion is changed little even if the raw landfill gas contains up to 60% methane, and clearly, as heating value requirement increases above 967 Btu/SCF, control of inerts in raw landfill gas feedstock becomes more crucial. (Heating value can be increased by propane injection. For example, 950 Btu gas could be boosted to 967 Btu gas by adding about 1 vol% propane; this would increase cost by about 15¢ per million Btu of product gas (propane @ 42¢/gallon, WSJ, 4/19/00).) The injection of processed landfill methane into distribution and transmission pipelines must be examined carefully on a case-by-case basis. There is precedent for co-mingling less than spec gas with pipeline gas, especially where demand is high and pipeline volumes large. For example, one large eastern gas utility upgrades subquality natural gas from about 300 Btu/SCF to 880 Btu/SCF for injection into its pipeline; the company's published acceptance heating value is 967 Btu/SCF. Little if any degradation of heating value occurs in the bulk mixed gas because of the relatively large volume of pipeline flow compared to the injected gas. Such exceptions might be granted for landfill methane, but lengthy negotiations with the pipeline owner can be anticipated.

B.2.1 BGE Miscellaneous Pipeline Data

Together with the few actual gas properties requirements provided by Baltimore Gas & Electric Company (BGE), several additional items of information which may be pertinent to future development of landfill gas to pipeline gas projects include:

- ***Typical pipeline pressure:*** BGE's "high pressure" lines range from 70 psig to 100 psig. BGE's "over high pressure" lines range between 100 psig and 300 psig.
- ***Estimated cost of connecting to a gas main:*** BGE responds with $800 to $9,500. This amount appears low compared to a gate station facility quoted by Columbia Transmission, order of magnitude several hundred thousand dollars. We will review this estimate with subcontractor Vandor. The difference may be pipeline pressure.
- ***Estimated cost of building a pipeline:*** $8 to $71 per foot of new pipeline, depending on pipe size and field conditions.

Table B.5

Natural Gas Specifications for Acceptance into Transmission or Distribution Pipelines

Property/Characteristic	Peoples	Columbia Transmission	Equitable	Baltimore Gas & Electric
Water, lb/MMSCF	5	7	7	—
CO2, vol%	3	1.25	3	—
H2S, grain/100 SCF **	¼	¼	0.3	—
Total Sulfur, grain/100 SCF	10	20	30	—
Heating Value, Btu/SCF	***1,000***	***967***	***1,000***	***1,000***
Utilization Factor*	1300	1300±6	—	—
Particulates	—	"free"	"free"	—
Liquids	—	"free"	"free"	—
Carbon Monoxide	—	—	0.1% vol	—
Inerts (CO2+N2+A+He)	—	***4% total***	***4% total***	—
Oxygen	—	0.02% vol	1% vol	—
Gasoline	—	—	0.2 gal/MCF	—
Temperature	—	—	< 100°F	—
HC Dewpoint	—	> 25°F	—	—

*Utilization Factor = heating value (Btu/SCF) / $\sqrt{\text{specific gravity}}$

**1 grain H2S/100 SCF is equivalent to 16 ppm by volume

B.3 Trading Landfill Methane for Remote LNG

Vandor + Vandor, independent subcontractor to Acrion, completed its preliminary study of "wheeling" landfill methane for LNG. Vandor's work examined the feasibility of utilizing excess liquefaction capacity at large peak shaving plants to produce LNG, and distribution of LNG from the peak shaving plant(s) within Maryland. Vandor's concept of swapping landfill methane for methane liquefied at remote large scale peak shaving plants is summarized in Vandor's letter final report attached as Appendix B. The concept appears feasible.

Table B.6
Gas Quality Task Force Survey
March 1995 [1]

Btu Content

min Btu/SCF	# pipelines	max Btu/SCF	# pipelines
930	1	950	1
950-960	18	1050-1100	9
961-970	17	1101-1150	2
971-980	3	1151-1200	5
981-1000	3	>1200	2
>1000	1	HC dewpoint limit	7
	43 tot		26 tot

950 Btu "wet" = 967 Btu "dry" basis, wet/dry distinction slowly lost over time

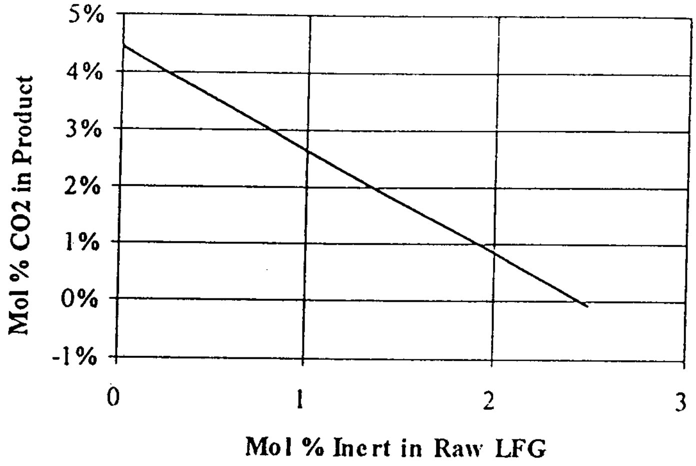
Sulfur Content

H2S max grain/100SCF	# pipelines	Total Sulfur grain/100 SCF	# pipelines
¼	23	½	1
0.3	3	¾	1
0.5	1	1	1
1	16	2	2
	43 tot	5	7
		10	4
		20	26
			42 tot

Oxygen (ppm)	# pipelines	*CO2 max vol %*	# pipelines
10	8	1	1
20	2	2	22
50	3	3	15
500	1		38 tot
1000	1	*Water, lb/MMSCF*	# pipelines
2000	11	4	8
2500	1	5	5
4000	6	6	2
10,000	6	7	27
	39 tot		42 tot

Figure B.2

Heating Value 967 Btu/SCF



C. Subtask 2.2
Sewage Digester Gas Process Flowsheet Development

C.1 Digester Gas as a Renewable Energy Source

C.1.1 Domestic Waste Water Treatment (WWT) Plants

In the USA, the volume of domestic waste water produced per capita ranges from 100 to 175 gallons per day, depending on location and season. Domestic waste water consists of organic and inorganic wastes. Organic wastes can be digested by bacteria and other microorganisms. Inorganic wastes are mineral substances such as sand, salt, iron, calcium, etc.; the latter are not digestible by microorganisms.

Sludge results from microorganisms having digested the organic portion of waste water; this is an exothermic reaction. Sludge needs to be stabilized to render it inactive and thus appropriate to be used for commercial applications such as compost. This stabilization process can be performed by anaerobic digestion (AD) which produces biogas, a mixture of methane, carbon dioxide, and other gases including nitrogen, oxygen and hydrogen sulfide.

Design flows of typical domestic WWT plants

WWT plants design flows range from several million gallons a day (MGD) to billions of gallons per day. Boston's new Deer Island facility now treats 1.3 billion gallon per day, and incorporates state of the art digesters which treat 50 ton per day of sludge. Waste water generation in a given metropolitan area can be estimated from the per capita amounts stated above. A metropolitan area such as Cleveland, Ohio, with several million inhabitants would be expected to generate several hundred million gallons per day of waste water requiring treatment.

Anaerobic digestion

In 1986, there were about 15,400 domestic municipal sewage treatment plants treating a combined flow of about 40 billion gallons per day. A major vendor of WWT equipment estimates that 25% of these WWT plants utilize anaerobic digestion technology to process the sludge produced. It is further estimated that of the 25% which utilize anaerobic digestion, 100 to 200 plants produce biogas in volumes greater than 2.5 million standard cubic feet per day (MMCFD).

Municipal anaerobic sludge digesters have recovered biogas for internal use for many years. Breaking into this integrated energy use within the WWT may be a barrier to upgrading the biogas for internal use or external sale. There has been an increased interest in controlled anaerobic digestion of municipal solid waste (MSW) as an alternative to landfills and incineration. In 1993, there were 15 plants in operation worldwide with 1 installation in the USA.

The treatment of municipal waste water results in the formation of slurries high in suspended solids. These slurries are commonly referred as to sludge. Sludge, an odorous, watery mixture; is the result of primary and secondary treatments that have removed solids, organic matter and bacteria from waste water; sludge needs to be stabilized before further use or disposal.

Anaerobic digestion of sludge includes two basic steps:

1) conversion of organic material to volatile acids, and
2) conversion of volatile acids to methane.

Bacteria that convert organic materials to volatile acids are called acid formers. Since acids are being formed, pH becomes very important. The anaerobic process requires pH to remain as close to 7.0 (neutral) as possible. Volatile acids are changed into methane by a second set of bacteria (methanogenic bacteria). Most digesters operate at 95°F (35°C) using mesophilic bacteria.

To maintain an anaerobic treatment system that will stabilize organic waste efficiently, the bacteria must be in a state of dynamic equilibrium; the reactor contents should be free of undesirable constituents and the pH of the aqueous environment should range from 6.5 to 7.5. The pH should not drop below 6.2 because the methane bacteria cannot function below this point. Other conditions involving nutrients and temperature must also occur.

The biodegradable portion of the organic fraction of MSW is converted biologically under anaerobic conditions to a gas containing carbon dioxide and methane (CH_4). The principal end products are carbon dioxide, methane, ammonia, hydrogen sulfide, and unreactive organic matter. In most anaerobic conversion processes carbon dioxide and methane constitute over 99% of the gas produced. Resistant organic matter (or digested sludge) must be dewatered before it can be disposed of by land spreading or landfilling. Dewatered sludge is often composted aerobically to stabilize it further before application.

Biogas usage

Some larger plants use sludge-produced biogas to power gas engines and generate electricity. In most case, smaller plants have insufficient gas to justify infrastructure for power generation, and simply dispose of sludge gas by flaring.

Biogas cleanup for engine fuel

Opinions as to the necessity of cleaning biogas before it is utilized to fuel internal combustion engines differ considerably. There is general consensus that water and hydrogen sulfide should be removed from biogas prior to combustion in the engine. Beyond these two obligations, other biogas contaminants removal processes are implemented within a wide range.

Engine manufacturers have developed specific fuel gas specifications limiting the contaminant content within the fuels burned.

<u>Carbon Dioxide</u>: affects engine performance and emissions in many ways. During combustion, water vapor and CO_2 can form carbonic acid that can attack certain engine parts and foul engine oil. However, it is generally not considered economical to remove CO_2 from engine fuel.

<u>Hydrogen Sulfide:</u> engine manufacturers recommend that H_2S be limited to under 10 ppm or 0.001% by volume. Caterpillar recommends that H_2S levels be lower than 47.5 µg/Btu.

<u>Volatile Organic Compounds:</u> LFG can also include volatile organic compounds (VOCs)

and chlorofluorocarbons (CFCs) which, when burned in an engine, form hydrochloric acid and hydrofluoric acid which can corrode engine components. Caterpillar recommends keeping chlorine and fluorine levels below 40 μg/Btu.

Water: water vapor combines with contaminants to form organic acids and carbonic acid which contaminate engine oil. Caterpillar recommends that moisture content be kept below 0.1 lb/MSCF.

Particulates: silica particulates should be kept below 0.4 microns to prevent abrasion. There is evidence that engine manufacturers are now designing contaminant-resistant engines; there is also evidence from the field that unexpectedly high maintenance costs occur when raw landfill gas (having undergone only a preliminary moisture removal process) is fed to engines. Information coming from Europe in the course of recovering biogas from collective organic waste treatment suggests that regular contracted engine maintenance helps prevent unexpected engine breakdowns.

It is generally recognized the all biogas conversion systems will require some form of gas cleanup. While the removal of water and H_2S are the only cleanup required for many applications, more extensive techniques may be required for more complex applications. H_2S content is usually less than 1%. Corrosion may occur however, especially if the biogas is compressed. Biogas can typically be combusted in a burner or engine without exceeding SO_2 emission limits. Typical H_2S levels in biogas are shown in Table C.1.

Table C.1
H_2S Levels in Biogas

SOURCE	H_2S Levels
Animal waste	100-1,000 ppmv
Industrial waste water	100-10,000 ppmv
Municipal sludge	1,000-20,000 ppmv
Landfill	100-20,000 ppmv

CO_2 reduces the heating value of biogas; removing CO_2 increases heating value. High Btu gases of pipeline quality require the removal of sulfides, CO_2 and water.

Gas Production in anaerobic digestion

Anaerobic processes producing methane are temperature sensitive; therefore temperature control is necessary. Most process are maintained at approximately 95°F (35°C) and must be heated. A portion of the methane produced is often used to heat the system. Gas production varies with temperature; it ranges from approximately 6.0 to 7.2 ft3 of gas per lb volatile solids added, or 7.0 to 15.1 ft3 gas per volatile solids digested. The heating value of raw sewage digester biogas is roughly 580 to 650 Btu/ft3.

Low-solids and high-solids anaerobic digestion

Low-solids anaerobic digestion is a biological process in which organic wastes are fermented at solids concentrations less than about 4 to 8 percent. This low-solid process is used in many parts of the world to generate methane gas from human, animal and agricultural waste as well as from the organic fraction of MSW. The process occurs in three steps: 1)receiving and preparation; 2) addition of moisture and nutrients, pH adjustment, heating; capture, storage and separation of the gas components CH_4 and CO_2; 3) dewatering and disposal of the digested sludge. Gas production is the range 8 to 16 ft3/lb (0.5 to 0.75 m3/kg) of volatile solids destroyed.

High-solid digestion is a newer technology; it is similar to low-solid digestion except that total solid content is about 22%. This process requires less water and produces higher gas volumes per unit waste digested. Beginning 1992, high-solid anaerobic digestion installations were operating in Europe, while several high-solid process were in development in the United States. As of 1992, neither low of high solids digestion technology had been commercialized for energy recovery.

Combustion of sludge

Multiple technologies are available for sludge combustion. One example is the multiple hearth incineration technology manufactured by Zimpro, Wheelabrator Incineration, C-E Raymond. This technology features the combustion of sludge through successive chambers with varying temperatures from 300°F to 1,800°F. This is the most widely used technology in the United States (350 installations). Waste heat is made available for power generation.

C.1.2 Characteristics of Sewage Digester Biogas

To investigate and record the characteristics of biogas produced by anaerobic digestion within WWT plants, AplusB, Inc, sent a questionnaire to 64 WWT plants servicing the largest metropolitan areas in Ohio. Replies totaled 6 in all; within these 6 replies, only one plant indicated the characteristics of the biogas produced; results are listed in Table C.2.

Table C.2
Municipal WWT Plant Biogas Characteristics

Characteristics	Plant data
Methane	65.74%
Carbon Dioxide	33.99%
Nitrogen	0.27%
Oxygen	0%
Water	6%
Hydrogen Sulfide	190 ppm
Carbonyl sulfide	0.23 ppm
Ethanol	0.16 ppm
Btu gross saturated	657
Btu gross dry	699

C.1.3 Summary Study of Digester Gas (Biogas)

AplusB, Inc., independent subcontractor to Acrion, completed its study of sewage digester gas, and a review of commercial or near-commercial technologies and companies prominently associated with bio-digesters. AplusB's digester gas study final report is attached as Appendix C.

At this time, Acrion does not intend to actively purse sewage digester gas as a viable source of methane for conversion to LNG heavy duty truck fuel for several reasons: 1) digester gas is already tightly integrated into the energy balance of most waste water treatment plants (WWTP), and conversion of digester methane to LNG for off-site use would require procurement of an alternative less expensive energy source to make economic sense and maintain energy balance within the WWTP; and 2) there is greater opportunity to develop productive end uses of LNG as motor fuel for refuse trucks at or near municipal landfill sites.

C.2 Sewage Digester Gas Process Flowsheet Development

The purpose of this task is to examine the use of sewage digester gas, a mixture of methane and carbon dioxide with small amounts of trace contaminants, as feedstock for liquid methane and carbon dioxide (CO_2) production. Acrion has developed a patented process (US Patent 5,681,360) which removes contaminants from CO_2 containing gas streams and produces a clean mixture of methane and CO_2 suitable for processing to a variety of finished products. Sewage digester gas differs from landfill gas on several accounts: it has a lower CO_2 content, and it typically contains more H_2S.

C.2.1 Capital and Operating Cost Estimates

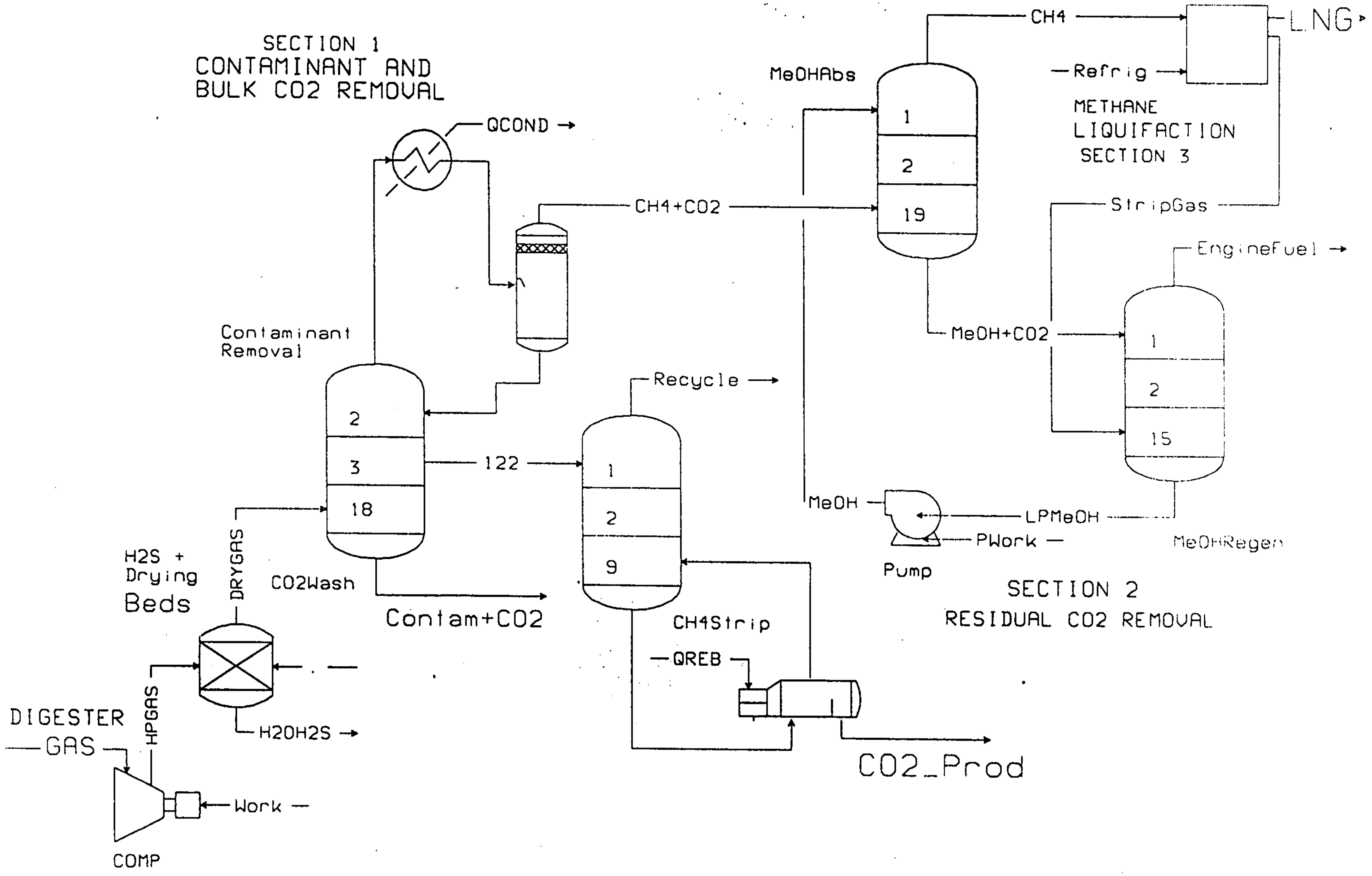
A process flowsheet, material balance, process energy requirements, equipment sizes, and capital and operating costs for the production of 19,000 gal/day of liquid methane and 54.8 tons/day of liquid CO_2 from 3.54 million standard cubic feet/day (MMSCFD) (dry basis) of sewage digester gas are presented. The assumed gas composition is 61.5% methane, 1.1% nitrogen and 37.4% carbon dioxide. Economics for H_2S levels of 100 and 600 ppm in the digester gas are presented. A process description is given below. A process flow diagram showing the major process sections is shown in Figure C.1. Detailed flowsheets followed by a material balance for each flowsheet are presented in Appendix D.

Process Description

The process comprises three main sections: 1) contaminant and bulk carbon dioxide removal, 2) residual carbon dioxide removal to 50 ppm, and 3) methane liquefaction. Hydrogen sulfide is removed in a separate adsorption bed up stream of the contaminant and bulk CO_2 removal section.

In Section 1, contaminant and bulk CO_2 removal, raw landfill gas is compressed, dried, and cooled to condense a majority of the CO_2. Process Section 1 is shown in Figure D-1, Appendix D. A portion of the liquid CO_2 condensate is used to absorb contaminants; the remainder is food-grade liquid CO_2 product.

Figure C.1
Process Flow Diagram
Digester Gas Conversion to Liquid Methane



Raw digester gas (stream 99) along with a recycle stream is compressed and cooled to condense water. The gas is passed through a hydrogen sulfide removal bed and a water removal bed. The hydrogen sulfide removal bed contains an iron based sorbent. Hydrogen sulfide reacts with the sorbent to produce iron sufide and water vapor. The sorbent is sold commercially under the trade names Sulfatreat (The Sulfatreat Company) and Sulfur Rite (U.S. Filter). Two beds are used with one on-line and the other off line. When H2S begins to breakthrough the on-line bed the beds are switched and spent sorbent is removed and replaced with fresh sorbent in the off-line bed. The spent material is non-hazardous. Depending on the H2S level the beds are switched every 2 to 6 months. In the water removal bed the gas is dried over a molecular sieve. Two beds are again used with bed switching every 8 hours. The water removal bed is regenerated by heating and purging. The dry gas (stream 109) is further compressed to 725 psia and cooled by heat exchange with product streams in HX104. The cold dry gas enters contaminant absorption column (abs 100) where it counter-currently contacts liquid carbon dioxide. Liquid CO2 absorbs all the contaminants from the gas.. The contaminant rich absorbent (stream 113) from the bottom of absorber is reduced in pressure and heated, by heat exchange with condensing CO2, in HX108 to vaporize a portion of the CO2. This vapor is warmed against the feed and recycled back to the inlet of the last compressor. The remaining liquid is vaporized and warmed to ambient temperature for cooling recovery and piped (stream 118) to the existing digester gas flare where contaminants are thermally oxidized.

Contaminant-free gas leaving the top of contaminant absorption column (abs 100) is further cooled in exchanger qcond to condense carbon dioxide. This condensate is separated from the gas stream and returned to the top of contaminant absorption column (abs 100) as absorbent. Liquid carbon dioxide does not freeze at this temperature due to the presence of dissolved methane in the liquid. The contaminant-free gas (stream 124) from this condensation step contains 15% carbon dioxide, and is sent to process section 2, final carbon dioxide removal (as stream 200).

A portion of condensate liquid CO2 (stream 122) is removed near the top of contaminant absorber (abs 100). Stream 122 is flashed to 250 psia and enters the top of methane stripper (strip). Dissolved methane and nitrogen are stripped from liquid CO2 with reboiled vapor. Bottoms from the light ends stripper (stream 125) is food grade liquid CO2 product. Vapor from the top of the stripper (stream 123) is combined with stream 206, vapor from Section 2 final CO2 removal section and stream 120, contaminant recycle vapor, and is warmed by heat exchange with feed gas and recycled to the last stage of feed compression.

In Section 2, shown in Figure D-2, Appendix D, residual CO2 is removed from a clean binary mixture of methane and CO2 by absorption in cold methanol; methane thus treated contains less than 50 ppm CO2 and is suitable for liquefaction. Spent methanol absorbent is depressurized in two flashes to remove dissolved CO2 and methane. A portion of the methanol absorbent is stripped of CO2 with methane rich gas obtained from the methane liquefaction section and is sent back to the top of the absorber.. The remaining absorbent is sent to the lower portion of the absorber to help reduce the temperature rise caused by the heat of absorption of the CO2. Gas from the high pressure flash is recycled to the last stage of LFG compression, and gas from the low pressure flash is recycled to the feed. Both gas streams are warmed by heat exchange with the high pressure LFG feed.

In Section 3, methane liquefaction, methane is condensed to produce liquid methane. Process Section 3 is shown in Figure D-3, Appendix D. Methane is condensed (HX302) with ethylene refrigeration. It is subcooled with cooling supplied from vaporized liquid and warmed recycle gas in HX303. A portion of the subcooled liquid is depressurized, vaporized to supply cooling to HX303 and recompressed and recycled to the methane condenser. The remaining liquid is cooled to near atmospheric pressure. There are two pressure reduction steps. A portion of the vapor from the first flash, which is enriched in nitrogen is used as stripping gas for the residual CO_2 removal step. Removing this gas prevents a buildup of nitrogen in the recycle loop. The remaining vapor and vapor from the second flash are recompressed and recycled to the methane condenser. These recycle vapor streams are warmed by heat exchange with the compressed recycle stream.

Process Energy Requirements

Power is required for digester gas compression, refrigeration and methane recompression and pumping. Refrigeration is needed in feed gas cooling, CO_2 liquefaction, and methane liquefaction. The ethylene and propane cascade refrigeration system flowsheets are shown in Figures D-4 and D-5, Appendix D, respectively. The heat exchange network for feed gas cooling and CO_2 condensation are shown in Figure D-6.

The energy requirements for digester gas, propane, ethylene and methane compression and absorbent pumping are shown in Table C.3. The total power requirement is 2212 hp or 1650 kW. This power can either be purchased or a portion of it can be generated onsite from the contaminant free vent gas from the methanol solvent regenerator.

Equipment Size

Equipment sizes are shown in Table D-7 in Appendix D. LNG and liquid CO_2 storage tanks were sized to store 3 day's production. Vessels and tower diameters were sized using the HYSIM process simulator. Heat transfer coefficients used in heat exchanger design were between 25 (low pressure gas) and 100 (boiling-condensing) Btu/hr/ft2/F°. Dehydration beds were designed for an 8 hour cycle time.

Process Economics

Tables C.4 and C.5 show estimated capital and annual operating costs and annual revenues for the process. Table C.4 shows costs for a digester gas with 100 ppm H_2S and Table C.5 shows costs for a gas with 600 ppm H_2S. In both tables most of the power is generated onsite using the contaminant free vent gas from the methanol solvent regenerator. Product prices were 40¢/gal for liquid methane and $40/ton for liquid CO_2. The simple payback period for the 100 ppm H_2S case is 3.0 years.. The simple payback with 600 ppm H_2S digester gas is 3.3 years.

Table C.3
Process Power Requirements

		hp	kW
Gas Compression			
	Stage 1	391	
	Stage 2	398	
	Stage 3	422	
	Total	1,211	903
Propane Refrigeration			
	Low Pressure	233	
	High Pressure	281	
	Total	514	383
Ethylene Refrigeration			
	Low Pressure	66	
	High Pressure	163	
	Total	229	171
Methane Recompression			
	Low Pressure	22	
	Mid Pressure	48	
	High Pressure	164	
	Total	234	175
Pumps			
	Methanol	9	
	Methanol	14	
	Carbon Dioxide	1	
	Total	24	18
Grand Total		**2,212**	**1,650**

Table C.4
Capital, Operating Costs, and Revenues
Digester Gas Conversion to LNG and Liquid CO2
(100 ppm H2S)

RAW GAS CONSUMPTION (dry)		3.54 MM SCFD	62% CH4
OPERATING PARAMETERS			
Methane Recovery		83%	
CO2 Recovery		44%	
Onstream Factor, 350 days/year		96%	
PRODUCT	PRICE		
RLM	40 ¢/gallon		
Liquid CO2	$40/ton		
Power Required	1643 kW		
Power Generated	1136 kW	from Clean LFG	
CAPITAL COSTS		Installed	
Equipment		Cost	
Compressors		3,103,000	
Storage Tanks		523,000	
Columns		904,000	
Vessels		509,000	
Dehydration		160,000	
Heat Exchangers		1,091,000	
H2S Removal (100 ppm H2S)		101,000	
Pumps		177,000	
Power Generation		643,000	
Capital Cost		7,211,000	
Contingency @ 18%		1,298,000	
Total Capital Costs		8,509,000	
ANNUAL OPERATING EXPENSES			
Electric Power	5.00 ¢/kWh	213,000	
SulfaTreat (100 ppm H2S)		33,000	
Labor ($12/hr, 2 operators/day)		67,000	
Labor Overhead (100% of labor)		67,000	
Maintenance Materials (2% of capital)		170,000	
Maintenance Labor (3% of capital)		255,000	
Taxes & Insurance (1.5% of capital)		128,000	
Total Operating Costs		933,000	
ANNUAL INCOME	Amount	Daily	Annual
Liquid Methane	21,500 GPD	8,600	3,010,000
CO2	54.8 TPD	2,191	767,000
Total Income		**10,791**	**3,777,000**
PAYBACK PERIOD	3.0 Years	ACRION TECHNOLOGIES	

Table C.5
Capital, Operating Costs, and Revenues
Digester Gas Conversion to LNG and Liquid CO2
(600 ppm H2S)

RAW GAS CONSUMPTION (dry)		3.54 MM SCFD	62% CH4
OPERATING PARAMETERS			
Methane Recovery		83%	
CO2 Recovery		44%	
Onstream Factor, 350 days/year		96%	
PRODUCT	PRICE		
RLM	40 ¢/gallon		
Liquid CO2	$40/ton		
Power Required	1643 kW		
Power Generated	1136 kW	from Clean LFG	
CAPITAL COSTS			
Equipment		Installed Cost	
Compressors		3,103,000	
Storage Tanks		523,000	
Columns		904,000	
Vessels		509,000	
Dehydration		160,000	
Heat Exchangers		1,091,000	
H2S Removal (600 ppm H_2S)		300,000	
Pumps		177,000	
Power Generation		643,000	
Capital Cost		7,410,000	
Contingency @ 18%		1,334,000	
Total Capital Costs		8,744,000	
ANNUAL OPERATING EXPENSES			
Electric Power	5.00 ¢/kWh	213,000	
SulfaTreat (600 ppm H2S)		198,000	
Labor ($12/hr, 2 operators/day)		67,000	
Labor Overhead (100% of labor)		67,000	
Maintenance Materials (2% of capital)		175,000	
Maintenance Labor (3% of capital)		262,000	
Taxes & Insurance (1.5% of capital)		131,000	
Total Operating Costs		1,113,000	
ANNUAL INCOME	Amount	Daily	Annual
Liquid Methane	21,500 GPD	8,600	3,010,000
CO2	54.8 TPD	2,191	767,000
Total Income		**10,791**	**3,777,000**
PAYBACK PERIOD	3.3 Years	ACRION TECHNOLOGIES	

C.3 Hydrogen Sulfide Removal

A salable CO_2 product (either from landfill gas or sewage digester gas) requires all hydrogen sulfide be removed. CO_2 wash is not efficient at H_2S removal due to the high volatility of H_2S in CO_2. The H_2S removal process must selectively remove H_2S in the presence of CO_2. Solid and liquid scavengers for H_2S removal are commercially available. Solid scavengers react H_2S with iron, zinc or copper oxide to produce iron, zinc or copper sulfides. Iron based scavengers include Sulfatreat (The Sulfatreat Company) and Sulfur-Rite (U.S. Filter). A zinc based scavenger, G-72E, is sold by United Catalysts Inc. Calgon Carbon makes a copper oxide impregnated activated carbon called Sulfasorb. Liquid based scavengers are triazine or chelated iron based. Triazines include Baker Petroline's HSW700, Coastal Chemical's Sulfaguard and Quaker Chemical's Enviro-Scrub and Enviro-Tek. Enviro-Scrub, Sulfaguard and HSW700 are single use, throwaway products. Enviro-Tek can be regenerated. LO-CAT is a regenerable chelated iron based scavenger sold by U.S. Filter.

The iron based solid scavengers, Sulfatreat and Sulfur-Rite, react H_2S with iron oxide to form iron sulfide and water vapor. Sulfatreat has been used commercially on compressed (120 psig) and atmospheric pressure landfill gas for H_2S removal.. The unit must be placed upstream of dehydration because it requires the presence of water vapor to work. It works best when the gas is water saturated. Approximately 8.5 lbs of Sulfatreat or Sulfur-Rite remove 1 lb of H_2S. Sulfatreat costs $0.35/lb including $.06/lb shipping and Sulfur-Rite costs $.36/lb. The cost per pound of H_2S is $3.

The size of the sulfur removal bed is dependent on change out time. Two vessels connected in series are used in a lead lag arrangement. When the first vessel (upstream vessel) reaches capacity it is taken off line and flow passes directly to the second vessel. The first vessel is recharged with fresh media and placed back on line downstream of the second vessel. This allows continuous operation and the media is used to its full capacity. Short term spikes of H_2S in the LFG can be handled easily due to the long change out times. The spent media is non-toxic and can be disposed of in the landfill.

Sulfasorb, a copper impregnated activated carbon, is expensive. Sulfasorb 12 costs $85/lb of H_2S (capacity 5.1% H_2S at 4.38$/lb). It can be regenerated with steam and air a limited number of times, but regeneration was not recommended at high H_2S concentrations.

Non-regenerable liquid H_2S triazene scavengers have a capacity range from 0.6 to 1.1 gal/lb of H_2S. Enviro-Scrub costs about $7/gal. This cost is nearly twice the solid scavenger cost on a per pound of H_2S basis.

Quaker Chemical's liquid scavenger, Enviro-Tek, an aqueous solution of a triazine, can be regenerated. The process has not been used commercially on LFG. The solvent contacts LFG in an absorption tower. LFG must be saturated with water at operating conditions or water must be added if it is not saturated. Spent solvent is sparged with air in a separate tank. Solvent regeneration produces solid sulfur which must be filtered from the solution. The liquid is pumped through the filter to a holding tank, and then pumped back to the top of the absorption tower. Sparger air must either be incinerated by using the it as the air supply to the contaminant flare or contacted with bleach (sodium hypochlorite) to remove odors. The filter must be cleaned daily. Byproduct from filter cleaning is a slurry of solid sulfur. Triazine solution is added to make-up for losses in filter cleaning. Power is required for air blowers and a pump. Other operating costs include make-up triazine, labor for daily filter cleaning and disposal costs for the sulfur slurry. The process has a higher capital cost but lower operating costs than solid scavengers.

LO-CAT (U.S. Filter) operates in a similar manner to Enviro-Tek, with solvent regeneration using air and sulfur removal from the regenerated solvent using a bag filter. The system uses a 3 stage absorber to achieve an outlet H_2S concentration of 4 ppm, with a Sulfur-Rite guard bed to remove the final amounts of H_2S

D. Subtask 2.3 Mixed Refrigerants for Process Cooling

The temperatures required for CO_2 and methane condensation are lower than can be achieved with a single refrigerant operating above atmospheric pressure. Two refrigerants are used in a cascade system. A mixed refrigerant is capable of reaching these low temperatures in a single system. Mixed refrigerants provide cooling by vaporizing over a range of temperatures. Two cases of the use of mixed refrigerants for process cooling have been studied, one for the production of liquid CO_2 and LNG and the other for the production of liquid CO_2 and pipeline gas.

D.1 Refrigeration for LNG and Liquid CO_2 Production

Process designs developed in Phase I to produce LNG and liquid CO_2 from landfill gas (LFG) used a cascade refrigeration system with two separate refrigerants, ethylene and propane. The use of a single mixed refrigerant [2] has been investigated for one of the process designs generated in Phase I. Case 2, (as described in Phase I Quarterly Progress Report 3) which produced 90 TPD of CO_2 and 21,500 gpd of LNG using methanol wash as the final CO_2 removal process was selected. Refrigeration for this case was required for feed cooling, CO_2 condensation and methane condensation.

A process flow diagram for the mixed refrigeration system is shown in Figure D.1. The material and energy balance for the flowsheet is given in Table D.1. Stream numbers for the process streams to be cooled match those in the Phase I design. The mixed refrigerant has a composition of 2% methane, 56% ethylene and 42% propane. The mixed refrigerant vapor is compressed from just above 1 atmosphere to 670 psig in three stages of compression (comp 1,2,3) with intercooling (air 1,2) between each stage. The vapor is condensed in an air cooler (air3). The condensed liquid (stream 10) is subcooled (cool0) from 100°F to 58°F. A portion of the liquid is expanded to 345 psig and is evaporated to provide cooling to both refrigeration subcooler cool0 and the process feed cooler. The vapor (streams 17,18) from both heat exchangers is recycled to the inlet of the last stage of compression. The remaining liquid (stream 12) is subcooled again (cool1) to -15°F. A portion of this liquid (stream 21) is expanded to 117 psig and is evaporated to provide cooling to first process CO_2 condenser (co2cond2) and then the refrigeration subcooler cool1. The vapor (stream 33) is recycled to the inlet of the second stage of compression. The remaining liquid (stream 20) is subcooled in cool2 and cool3 to -117°F and is expanded to near atmospheric pressure (5 psig). The refrigerant warms as it evaporates and provides cooling to the following heat exchangers: 1) process methane condenser (HX302), 2) low temperature refrigerant subcooler and process methane cooler (cool3), 3) intermediate temperature refrigerant subcooler (cool2) and 4) the low temperature process CO_2 condenser (co2cond). The closest temperature approach of 4°F is in the methane condenser. The heat release curve for methane cooling and condensing (HX302 and cool3) is shown in Figure D.2.

Figure D.1
Mixed Refrigeration System Process Flow Diagram
LFG to Liquid Methane and CO_2 (Phase I, Case 2)

Refrigerant Composition: CH_4 2%, C_2H_4 56%, C_3H_8 42%

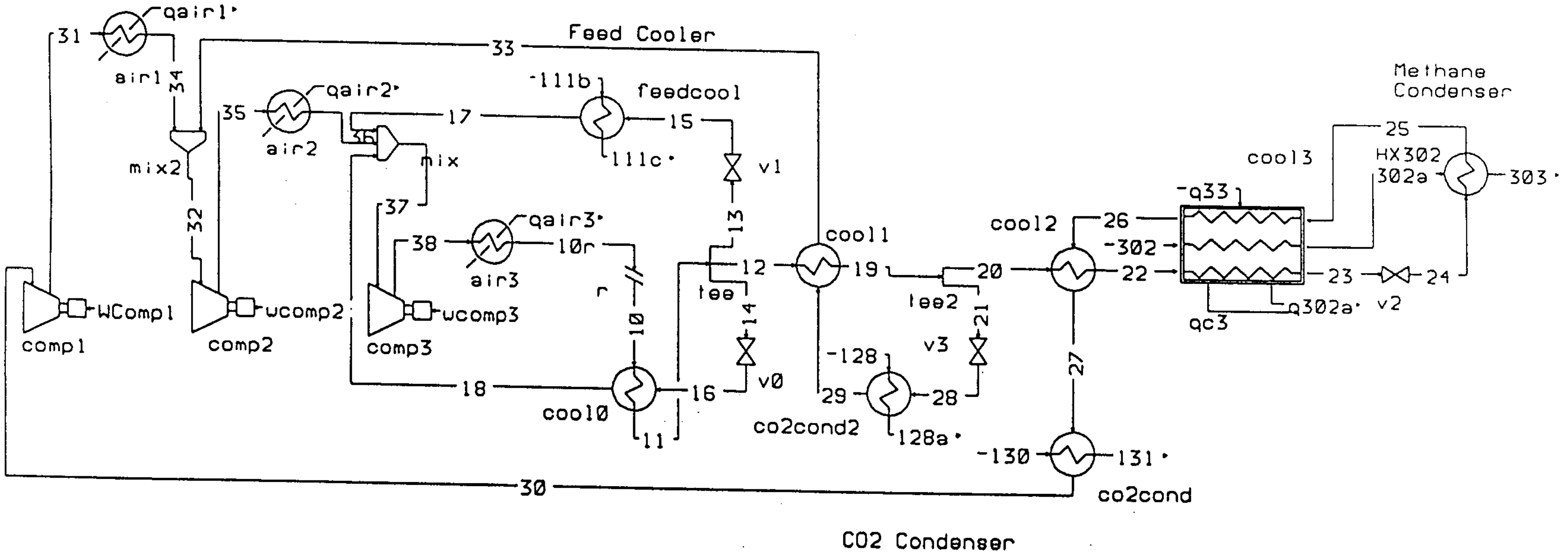


Table D.1
Mixed Refrigeration System Material and Energy Balance
LFG to Liquid Methane and CO2 (Phase I, Case 2)

Stream		111b	111c	128	128a
Vapour frac.		1.0000	1.0000	1.0000	0.9122
Temperature	F	100.0000*	58.4530*	-15.2605*	-21.4586*
Pressure	psia	715.0000*	710.0000*	700.8824*	700.0000*
Molar Flow	lbmole/hr	359.8113*	359.8113	694.5520*	694.5520
Enthalpy	Btu/hr	1.38390E+06	1.21397E+06	1.77521E+06	1.48480E+06

Stream		130	131	302	302a
Vapour frac.		0.7290	0.4407	1.0000	1.0000
Temperature	F	-38.6688*	-75.0000*	-85.0706*	-117.0000
Pressure	psia	700.0000*	700.0000*	680.0000*	675.0000*
Molar Flow	lbmole/hr	694.5520*	694.5520	442.7797*	442.7797
Enthalpy	Btu/hr	836775.9072	-146708.8674	963883.3666	651094.5144

Stream		303	10	11	12
Vapour frac.		0.0000	0.0000*	0.0000	0.0000
Temperature	F	-124.7006*	100.0000*	58.4530*	58.4530
Pressure	psia	670.0000*	680.6298	675.6298	675.6298
Molar Flow	lbmole/hr	442.7797	770.4043	770.4043	497.0100
Enthalpy	Btu/hr	212598.4486	1.13500E+06	50772.9372	32755.0848

Stream		13	14	15	16
Vapour frac.		0.0000	0.0000	0.1260	0.1260
Temperature	F	58.4530	58.4530	38.9133	38.9133
Pressure	psia	675.6298	675.6298	359.3771	359.3771
Molar Flow	lbmole/hr	37.0443	236.3500	37.0443	236.3500*
Enthalpy	Btu/hr	2441.3750	15576.4757	2441.3750	15576.4757

Stream		17	18	19	20
Vapour frac.		1.0000*	1.0000	0.0000	0.0000
Temperature	F	85.0000*	85.0087	-15.2605*	-15.2605
Pressure	psia	354.3771	354.3771	670.6298	670.6298
Molar Flow	lbmole/hr	37.0443	236.3500	497.0100	314.7900
Enthalpy	Btu/hr	172371.8352	1.09980E+06	-842051.2574	-533327.9379

Stream		21	22	23	24
Vapour frac.		0.0000	0.0000	0.0000	0.0673
Temperature	F	-15.2605	-85.0706*	-117.0000*	-132.5347
Pressure	psia	670.6298	665.6298	660.6299	20.0000*
Molar Flow	lbmole/hr	182.2200	314.7900	314.7900	314.7900
Enthalpy	Btu/hr	-308723.3195	-983749.7231	-1.17315E+06	-1.17315E+06

Table D.1 Continued
Mixed Refrigeration System Material and Energy Balance
LFG to Liquid Methane and CO2 (Phase I, Case 2)

Stream		25	26	27	28
Vapour frac.		0.2765	0.4965	0.6672	0.1099
Temperature	F	-121.0264	-104.8146	-89.1510	-37.5464
Pressure	psia	19.0000	18.0000	17.0000	131.6764
Molar Flow	lbmole/hr	314.7900	314.7900	314.7900	182.2200
Enthalpy	Btu/hr	-734655.5869	-232464.8043	217956.9809	-308723.3195

Stream		29	30	31	32
Vapour frac.		0.3703	1.0000	1.0000	1.0000
Temperature	F	-22.0000*	-48.6691	158.6650	79.6033
Pressure	psia	126.6764	16.0000	126.6764	121.6764
Molar Flow	lbmole/hr	182.2200*	314.7900*	314.7900	497.0100
Enthalpy	Btu/hr	-18307.9813	1.20144E+06	2.01027E+06	2.59116E+06

Stream		33	34	35	36
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	43.4472	100.0000*	202.9878	100.0000*
Pressure	psia	121.6764	121.6764	359.3771	354.3771
Molar Flow	lbmole/hr	182.2200	314.7900	497.0100	497.0100
Enthalpy	Btu/hr	856498.3732	1.73466E+06	3.33094E+06	2.44744E+06

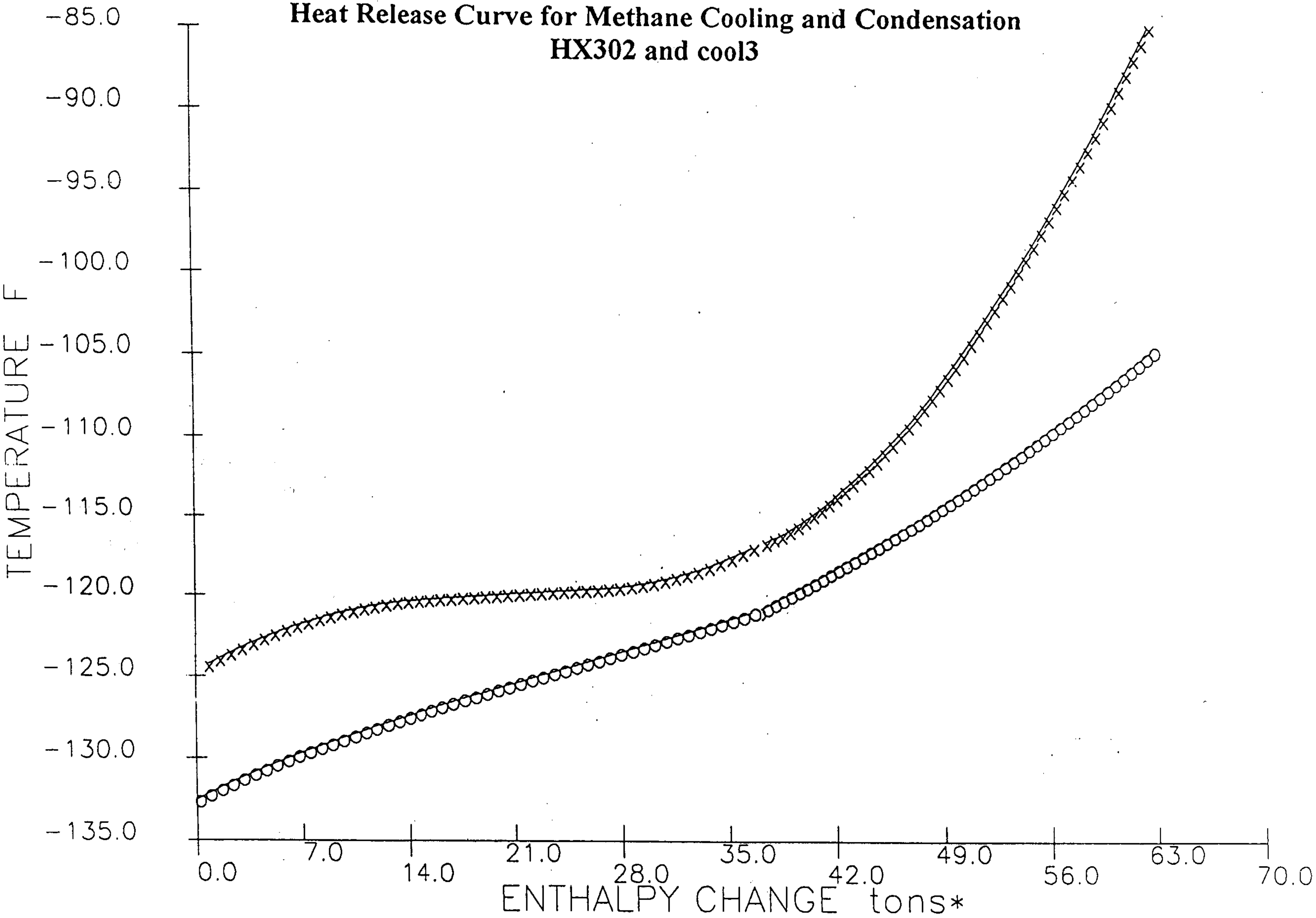
Stream		37	38	10r
Vapour frac.		1.0000	1.0000	0.0000
Temperature	F	94.6281	178.8808	100.0000*
Pressure	psia	354.3771	685.6298	680.6298
Molar Flow	lbmole/hr	770.4043	770.4043	770.4043
Enthalpy	Btu/hr	3.71962E+06	4.27707E+06	1.13500E+06

Stream		qair1	qair2	qair3	qc3
Enthalpy	Btu/hr	275610.5522	883502.5473	3.14207E+06	189401.9253

Stream		q302a	q33
Enthalpy	Btu/hr	312788.8478	502190.7731

Stream		WComp1	wcomp2	wcomp3
Enthalpy	hp	317.8831	290.7473	219.0855

Figure D.2
Heat Release Curve for Methane Cooling and Condensation
HX302 and cool3



The total power requirement for this design is 828 hp. This is only 0.4% (825 hp) higher than the cascade refrigeration case. One advantage over the cascade refrigeration case is that less compressors are required. The heat exchanger duties in Btu/hr, the weighted log mean temperature difference in °F and estimated areas in square feet are given in Table D.2. In calculating the heat exchanger areas, a heat transfer coefficient of 100 Btu/hr/ft^2/°F (for condensation and liquid cooling) was used in all exchangers except the feed cooler and methane cooler where as value of 70 Btu/hr/ft^2/°F (for gas cooling) was used.

Table D.2
Heat Exchanger Sizing

Heat Exchanger	Duty in Btu/hr	ΔT, °F	Area, ft^2
HX302	438,500	5.47	802
Cool3 (CH_4 cool)	312,800	7.20	621
Cool3 (subcool)	189,400	10.5	180
Cool2	450,400	41.5	109
CO2Cond	983,500	19.8	496
CO2Cond2	290,400	10.9	267
Cool1	874,800	16.2	541
Feedcool	169,900	15.1	161
Cool0	1,084200	17.3	626

Table D.3 compares the capital cost of the mixed refrigerant system to the cascade system. The mixed refrigerant system has a 5% lower capital cost with a lower compressor cost and a slightly higher heat exchanger cost.

D.2 Refrigeration for Pipeline Gas and Liquid CO_2 Production

The process design to produce pipeline gas and liquid CO_2 from landfill gas (LFG) described in section B.1 used a cascade refrigeration system with two separate refrigerants. Carbon dioxide was used as the low temperature refrigerant to provide cooling for CO_2 condensation from the LFG. The CO_2 refrigerant vapor was compressed to 300 psi and condensed at 0°F with the high temperature refrigerant. The use of a single mixed refrigerant has been investigated for this process design. The mixed refrigerant has a composition of 30% ethane and 70% propane.

Table D.3
Equipment Cost Comparison
Refrigeration for LFG to Liquid Methane and CO2

Mixed Refrigeration

Tag	Service	Size	Mat. + Press.	Equipment Cost, $K	Installed Cost K$
Compressors					
Comp1	Refrig Comp 1	318 hp		$192	$412
Comp2	Refrig Comp 2	291 hp		$182	$391
Comp3	Refrig Comp 3	219 hp		$154	$331
		828 hp		$528	$1,135
Exchangers					
FeedCool+Cool0	Feed Cooler + Refrig Subcooler 0	787 ft^2	CS	$23	$74
Cool1	Refrig Subcooler 1	541 ft^2	CS	$18	$57
Co2cond2	CO2 Cond -20F	267 ft^2	Al	$25	$79
Co2cond	CO2 Cond -70F	496 ft^2	Al	$38	$121
Cool2	Refrig Subcooler 2	109 ft^2	Al	$13	$42
Cool3	LNG Cool + Refrig Subcooler 3	801 ft^2	Al	$53	$170
HX302	LNG Cond	802 ft^2	Al	$54	$170
Air3	Refrigerant Condenser	2918.2 ft^2	CS	$14	$29
					$742
				Total	$1,877

Cascade Refrigeration
Ethylene(C2=) and Propane(C3)

Tag	Service	Size	Mat. + Press.	Equipment Cost, $K	Installed Cost K$
Compressors					
C2=CompL	Ethylene Stage 1	66 hp		$78	$167
C2=CompH	Ethylene Stage 2	167 hp		$132	$283
C3CompL	Propane Stage 1	268 hp		$173	$373
C3CompH	Propane Stage 2	324 hp		$194	$417
		825 hp		$577	$1,241
Exchangers					
FeedCool	Feed Cooler	54 ft^2	CS	$4	$11
Co2cond2	CO2 + C2= Cond -20F	2095 ft^2	Al	$105	$332
Co2cond	CO2 Cond -70F	553 ft^2	Al	$41	$131
HX302	LNG Cond	885 ft^2	Al	$57	$182
	C3 Condenser	10745 ft^2	CS	$35	$76
					$732
				Total	$1,973

CS= Carbon Steel
SS= Stainless Steel
KCS = Killed Carbon Steel
Al = Aluminum

A process flow diagram for the mixed refrigeration system is shown in Figure D.3. The material and energy balance for the flowsheet is given in Table D.4. Stream numbers for the process streams to be cooled match those in Figure B.1 and Appendix A. The mixed refrigerant vapor is compressed from just above 1 atmosphere (4 psig) to 318 psig in two stages of compression (Rcomp 1,2) with intercooling (air 1) between the stages. The vapor is condensed in an air cooler (air2). The condensed liquid (stream 2) is subcooled (HX7) from 100°F to 50°F. A portion of the liquid is expanded to 126 psig and is evaporated to provide cooling to refrigeration subcooler HX7. The vapor (stream 16) is recycled to the inlet of the last stage of compression. The remaining liquid (stream 4) is subcooled again (HX8, HX4) to -27°F and is expanded to near atmospheric pressure (6 psig). The refrigerant warms as it evaporates and provides cooling to the process CO_2 condenser (HX6). The closest temperature approach of 10°F is at the cold end of the CO_2 condenser. The refrigerant vapor is warmed to 2°F to provide some of the cooling in subcooler HX8. Cooling for subcooler HX8 is also provided by warming the liquid CO_2 product to its storage temperature of 0°F. Since the CO_2 is not compressed, the liquid needs to be pumped from the methane stripper pressure of 180 psia to the storage pressure of 300 psia. Cooling for subcooler HX4 is provided by vaporizing process contaminant rich CO_2. In the cascade design this cooling was used to condense refrigerant CO_2.

The total power requirement for the mixed refrigerant design is 318 hp. The cascade design in section B.1 (Table B.3) has a CO_2 compressor power requirement of 210 hp and the high temperature refrigerator horsepower is estimated at 243 hp to produce 94 Tons of cooling (twice the thermodynamic reversible work of a Carnot cycle operating between -20°F and 100°F). To better estimate the high temperature refrigerator power requirement a two stage ammonia refrigeration system has been designed operating at a -15°F evaporator temperature with a 55°F economizer. The power requirement for this refrigerator is 183 hp. The total combined power requirement for the CO_2 and ammonia cascade system is 393 hp. Therefore, the mixed refrigerant design has a 19% lower power requirement. An additional advantage over cascade refrigeration is fewer compressors are required.

The heat exchanger duties in Btu/hr, the weighted log mean temperature difference in °F and estimated areas in square feet are given in Table D.5. In calculating the heat exchanger areas, a heat transfer coefficient of 100 $Btu/hr/ft^2/°F$ (for condensation and liquid cooling) was used in all exchangers except HX8a where a value of 20 $Btu/hr/ft^2/°F$ (for low pressure gas heating) was used.

Table D.6 compares the capital cost of the mixed refrigerant system to the cascade system. The mixed refrigerant system has a 9% lower capital cost with a lower compressor cost and a slightly higher heat exchanger cost.

Figure D.3
Mixed Refrigeration System Process Flow Diagram
Pipeline Gas and CO2 Case

Refrigerant Composition: Ethane 30%, Propane 70%

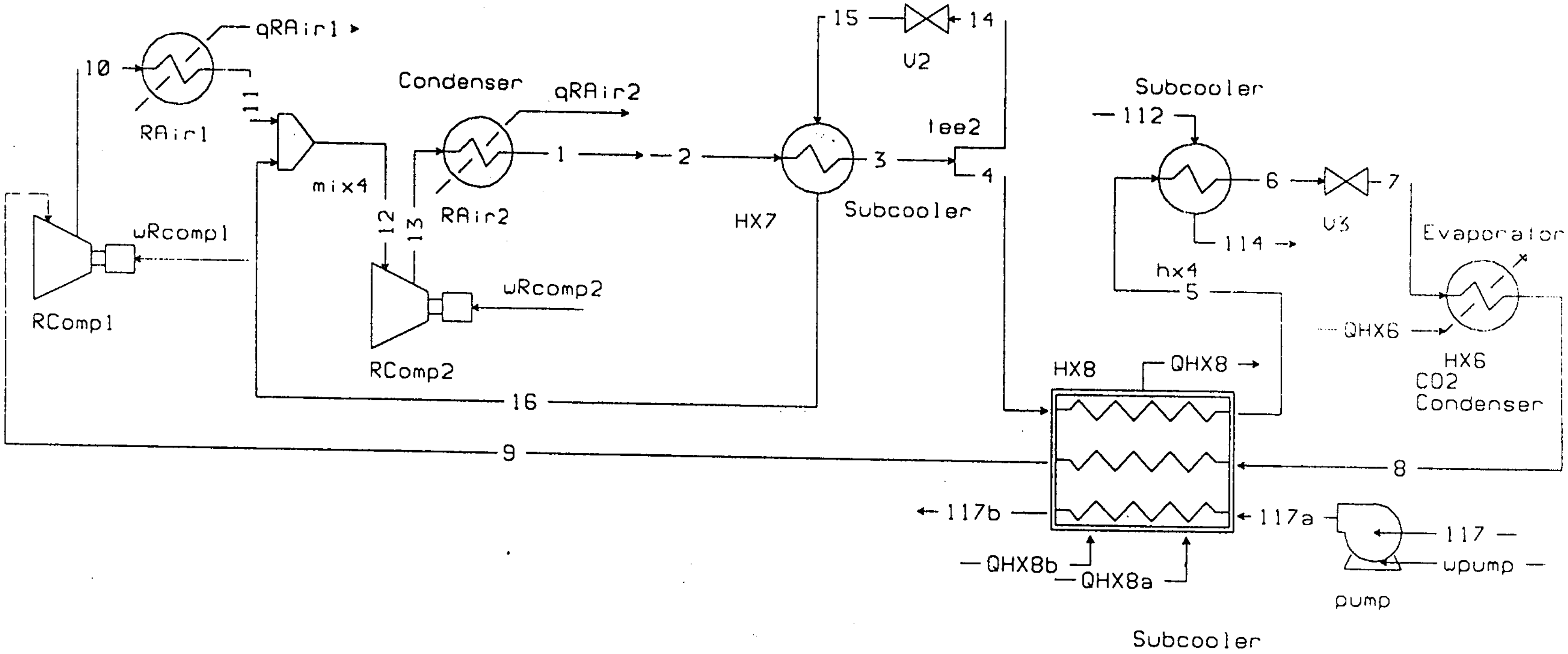


Table D.4
Material Balance for Mixed Refrigerant Flowsheet

Stream		1	2	3	4
Vapour frac.		0.0000	0.0000*	0.0000	0.0000
Temperature	F	100.0000*	100.0000*	50.0000*	50.0000
Pressure	psia	327.6762	327.6762	322.6762	322.6762
Molar Flow	lbmole/hr	246.7686	246.7686*	246.7686	191.4352
Enthalpy	Btu/hr	182938.4022	182938.4022	-186623.7428	-144776.7014

Stream		5	6	7	8
Vapour frac.		0.0000	0.0000	0.1390	1.0000
Temperature	F	7.0000*	-26.7763	-67.0000*	-33.3000*
Pressure	psia	317.6762	312.6761	20.7811	19.7811
Molar Flow	lbmole/hr	191.4352	191.4352	191.4352	191.4352
Enthalpy	Btu/hr	-357753.3007	-511378.4426	-511378.4426	853471.5634

Stream		9	10	11	12
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	1.9121	173.9582	100.0000*	96.4304
Pressure	psia	18.7811	135.0000*	130.0000	130.0000
Molar Flow	lbmole/hr	191.4352	191.4352	191.4352	246.7686
Enthalpy	Btu/hr	951836.5833	1.45023E+06	1.18867E+06	1.51639E+06

Stream		13	14	15	16
Vapour frac.		1.0000	0.0000	0.0717	1.0000
Temperature	F	189.8758	50.0000	35.0000*	85.0000*
Pressure	psia	332.6762	322.6762	140.3217	135.3217
Molar Flow	lbmole/hr	246.7686	55.3335	55.3335	55.3335
Enthalpy	Btu/hr	1.82061E+06	-41847.0414	-41847.0414	327715.1019

Stream		112	114	117	117a
Vapour frac.		0.0000*	1.0000	0.0000*	0.0000
Temperature	F	-33.9754	-28.1000*	-29.1238	-27.8348
Pressure	psia	180.0000*	175.0000	180.0000*	305.0000
Molar Flow	lbmole/hr	24.8871*	24.8871	167.2458	167.2458*
Enthalpy	Btu/hr	-80054.3608	73570.7847	-523300.5228	-518296.9469

Stream		117b	QHX8	QHX8a	QHX8b
Vapour frac.		0.0000*	2.0000*	2.0000*	2.0000*
Temperature	F	-1.4325	0.0000*	0.0000*	0.0000*
Pressure	psia	300.0000*	0.0000*	0.0000*	0.0000*
Molar Flow	lbmole/hr	167.2458	0.0000*	0.0000*	0.0000*
Enthalpy	Btu/hr	-403685.3736	212976.6048	98365.0229	114611.5746

Stream		QHX4	QHX6	qRAir1	qRAir2
Enthalpy	Btu/hr	153664.9926*	1.36485E+06	261556.7675	1.63767E+06

Stream		wRcomp1	wRcomp2	wpump
Enthalpy	hp	195.8780	119.5618	1.9665

Table D.5
Heat Exchanger Sizing

Heat Exchanger	Duty, Btu/hr	Temperature difference, °F	Area, ft^2
HX7	370,000	24.4	152
HX4	153,000	16.0	96
HX6	1,361,000	13.1	1039
HX8a	98,000	44.3	111
HX8b	115,000	42.8	27

Table D.6
Equipment Cost Comparison
Pipeline Gas Refrigeration System

Mixed Refrigeration (Ethane 30%, Propane 70%)

Tag	Service	Size		Mat. + Press.	Equipment Cost, $K	Installed Cost K$
Compressors						
Comp1	Refrig Comp 1	196	hp		$145	$311
Comp2	Refrig Comp 2	120	hp		$109	$235
		316	hp		$254	$545
Pump	CO2 Pump (with spare)	2	hp	SS	$15	$49
Exchangers						
HX4	Refrig Subcooler -Contam CO2 Eva	96	ft^2	SS	$15	$46
HX6	CO2 Condenser - Refrig Evap.	1039	ft^2	Al	$64	$203
HX7	Refrig Subcooler	152	ft^2	CS	$7	$24
HX8	Refrig Subcooler	138	ft^2	Al	$16	$50
Air2	Refrigerant Condenser	1171	ft^2	CS	$7	$15
						$338
					Total	$932

Cascade Refrigeration
CO2 and Ammonia

Tag	Service	Size		Mat. + Press.	Equipment Cost, $K	Installed Cost K$
Compressors						
CO2	CO2	210	hp		$150	$324
	NH3 Stage 1	114	hp		$106	$228
	NH3 Stage 2	69	hp		$80	$172
		393	hp		$336	$723
Exchangers						
HX4	CO2 Ref Cond - Contam CO2 E	55	ft^2	SS	$10	$31
HX6	CO2 Ref Evap- CO2 Cond	548	ft^2	Al	$41	$130
HX7-8	CO2 Ref heater-cooler	155	ft^2	Al	$17	$54
HX9	CO2 Ref Cond - NH3 Evap	781	ft^2	CS	$23	$74
	NH3 Condenser	1134	ft^2	CS	$7	$15
						$304
					Total	$1,027

CS= Carbon Steel
SS= Stainless Steel
KCS = Killed Carbon Steel
Al = Aluminum

E. Tasks 3 and 4 Demonstration Unit and Market Site

Waste Management's Arden Landfill near Washington PA is an excellent candidate landfill for demonstrating the production of LNG for heavy vehicle use from LFG. The site has a 13,000 Gal LNG tank which fuels 7 refuse trucks with natural gas engines. Acrion has prepared a cost estimate for a facility to supply clean methane to a 1200 gal/day liquefaction unit located at Arden. The unit would produce a 400 psia gas stream with less than 50 ppm CO_2 and less than 0.1 ppm water and no contaminants. No liquid CO_2 would be co-produced. The process steps are 1) compression 2) drying and Acrion CO_2 wash contaminant and bulk CO_2 removal, 3) additional CO_2 removal with a membrane and thermal swing CO_2 absorption (TSA) bed. Vendor quotes were obtained from the Wittimann Co. for the drying and CO_2 wash step and from UOP for the membrane and TSA unit. Wittimann built Acrion's pilot unit for the New Jersey Ecocomplex. The details of the cost estimate are in a letter report in Appendix E. Over 60% of the costs were for the UOP membrane and TSA unit. The estimated installed plant cost is $860,000.

Alternate vendors and methods of final CO_2 removal were also investigated. Instead of the membrane and TSA system an amine treater followed by an additional dryer or a vacuum pressure swing absorption (VPSA) unit could be used. Neither case gave better economics at this scale. A cost estimate of $250,000 for a membrane from Cynara (now called Natco) was obtained. The cost was for the smallest membrane they produce which had a capacity of nearly 10 times what was needed. It is clear from discussions with vendors that there would be large economies of scale for a larger plant.

Acrion's pilot LFG to liquid CO_2 pilot unit, currently under construction for placement at the NJ Ecocomplex, can supply high pressure contaminant free methane enriched gas which can be further processed to approximately 800 gal/day of LNG. The unit produces a gas stream at 400 psig containing 26% CO_2 with no contaminants or water. The unit also produces 1.8 tons/day of commercial liquid CO_2. Testing at the Ecocomplex will be completed in July 2001 and the unit is not currently scheduled for any other use.

The next step in the development process is to define the size, scope and funding for a project at the Arden landfill. Acrion would pursue the task of firming the cost estimate and identifying lower cost technologies and sources for trace CO_2 removal. We would also identify the market for co-produced CO_2 in the area.

F. References

1) "Grid Integration Project Gas Quality Task Force," Roger Huffaker, Gas Research Institute Report, March 1995.

2) Gas Processors Suppliers Association Engineering Data Book, Volume I, Tenth Edition, 1987, Page 14-22.

APPENDIX A

Acrion CO2 Wash + Selexol to Produce Pipeline Gas from LFG

Hysim Process Simulator Material and Energy Balance and Process Flowsheets

Acrion CO2 Wash + SELEXOL to produce CO2 + Pipline Gas from LFG

Stream Description		100 Feed	110 PipelineGas	117 CO2 Product
Methane	lbmole/hr	237.1744*	237.0622	0.0002
Nitrogen	lbmole/hr	4.3921*	4.3912	0.0000
CO2	lbmole/hr	197.6453*	5.5601	167.2442
SELEXOL	lbmole/hr	0.0000*	0.0000	0.0000
M-Mercaptan	lbmole/hr	0.0004*	0.0000	0.0000
Propane	lbmole/hr	0.0004*	0.0000	0.0004
Refrig-12	lbmole/hr	0.0004*	0.0000	0.0000
Refrig-40	lbmole/hr	0.0004*	0.0000	0.0000
VinylCl	lbmole/hr	0.0004*	0.0000	0.0000
COS	lbmole/hr	0.0004*	0.0000	0.0003
H2S	lbmole/hr	0.0004*	0.0000	0.0004
Toluene	lbmole/hr	0.0004*	0.0000	0.0000
Cl3-C2=	lbmole/hr	0.0004*	0.0000	0.0000
Refrig-22	lbmole/hr	0.0004*	0.0000	0.0000
n-Butane	lbmole/hr	0.0004*	0.0000	0.0000
n-Pentane	lbmole/hr	0.0004*	0.0000	0.0000
i-Pentane	lbmole/hr	0.0004*	0.0000	0.0000
22-Mpropane	lbmole/hr	0.0004*	0.0000	0.0000
i-Butane	lbmole/hr	0.0004*	0.0000	0.0002
Ethane	lbmole/hr	0.0004*	0.0003	0.0001
H2O	lbmole/hr	11.5000*	0.0000	0.0000
Total:	lbmole/hr	450.7188*	247.0138	167.2458

Stream Description		100 Feed	110 PipelineGas	117 CO2 Product
Methane	mole frac.	0.526213586*	0.959712267	0.000001000
Nitrogen	mole frac.	0.009744696*	0.017777236	0.000000000
CO2	mole frac.	0.438511342*	0.022509076	0.999990821
SELEXOL	mole frac.	0.000000000*	0.000000000	0.000000000
M-Mercaptan	mole frac.	0.000000974*	0.000000000	0.000000000
Propane	mole frac.	0.000000974*	0.000000047	0.000002173
Refrig-12	mole frac.	0.000000974*	0.000000000	0.000000122
Refrig-40	mole frac.	0.000000974*	0.000000000	0.000000001
VinylCl	mole frac.	0.000000974*	0.000000000	0.000000000
COS	mole frac.	0.000000974*	0.000000005	0.000001524
H2S	mole frac.	0.000000974*	0.000000024	0.000002116
Toluene	mole frac.	0.000000974*	0.000000000	0.000000000
Cl3-C2=	mole frac.	0.000000974*	0.000000000	0.000000000
Refrig-22	mole frac.	0.000000974*	0.000000000	0.000000000
n-Butane	mole frac.	0.000000974*	0.000000000	0.000000286
n-Pentane	mole frac.	0.000000974*	0.000000000	0.000000000
i-Pentane	mole frac.	0.000000974*	0.000000000	0.000000000
22-Mpropane	mole frac.	0.000000974*	0.000000000	0.000000002
i-Butane	mole frac.	0.000000974*	0.000000005	0.000001431
Ethane	mole frac.	0.000000974*	0.000001363	0.000000517
H2O	mole frac.	0.025514800*	0.000000000	0.000000000

Acrion CO2 Wash + SELEXOL to produce CO2 + Pipline Gas from LFG

Stream Description		100 Feed	110 PipelineGas	117 CO2 Product
Methane	MMSCFD	2.1600*	2.1589	0.0000
Nitrogen	MMSCFD	0.0400*	0.0400	0.0000
CO2	MMSCFD	1.8000*	0.0506	1.5231
SELEXOL	MMSCFD	0.0000*	0.0000	0.0000
M-Mercaptan	MMSCFD	0.0000*	0.0000	0.0000
Propane	MMSCFD	0.0000*	0.0000	0.0000
Refrig-12	MMSCFD	0.0000*	0.0000	0.0000
Refrig-40	MMSCFD	0.0000*	0.0000	0.0000
VinylCl	MMSCFD	0.0000*	0.0000	0.0000
COS	MMSCFD	0.0000*	0.0000	0.0000
H2S	MMSCFD	0.0000*	0.0000	0.0000
Toluene	MMSCFD	0.0000*	0.0000	0.0000
Cl3-C2=	MMSCFD	0.0000*	0.0000	0.0000
Refrig-22	MMSCFD	0.0000*	0.0000	0.0000
n-Butane	MMSCFD	0.0000*	0.0000	0.0000
n-Pentane	MMSCFD	0.0000*	0.0000	0.0000
i-Pentane	MMSCFD	0.0000*	0.0000	0.0000
22-Mpropane	MMSCFD	0.0000*	0.0000	0.0000
i-Butane	MMSCFD	0.0000*	0.0000	0.0000
Ethane	MMSCFD	0.0000*	0.0000	0.0000
H2O	MMSCFD	0.1047*	0.0000	0.0000
Total:	MMSCFD	4.1047*	2.2496	1.5231

Acrion CO2 Wash + SELEXOL to produce CO2 + Pipline Gas from LFG

Stream		101	106	107	111
Methane	lbmole/hr	283.1235	16.2723	258.5413	8.3099
Nitrogen	lbmole/hr	4.6559	0.1337	4.4530	0.0692
CO2	lbmole/hr	430.8012	223.5922	91.5928	115.6161
SELEXOL	lbmole/hr	0.0000	0.0000	0.0000	0.0000
M-Mercaptan	lbmole/hr	0.0005	0.0000	0.0000	0.0005
Propane	lbmole/hr	0.0009	0.0005	0.0002	0.0003
Refrig-12	lbmole/hr	0.0008	0.0000	0.0000	0.0008
Refrig-40	lbmole/hr	0.0007	0.0000	0.0000	0.0007
VinylCl	lbmole/hr	0.0006	0.0000	0.0000	0.0006
COS	lbmole/hr	0.0008	0.0003	0.0000	0.0005
H2S	lbmole/hr	0.0008	0.0004	0.0001	0.0003
Toluene	lbmole/hr	0.0004	0.0000	0.0000	0.0004
Cl3-C2=	lbmole/hr	0.0004	0.0000	0.0000	0.0004
Refrig-22	lbmole/hr	0.0006	0.0000	0.0000	0.0006
n-Butane	lbmole/hr	0.0008	0.0001	0.0000	0.0008
n-Pentane	lbmole/hr	0.0006	0.0000	0.0000	0.0006
i-Pentane	lbmole/hr	0.0006	0.0000	0.0000	0.0006
22-Mpropane	lbmole/hr	0.0007	0.0000	0.0000	0.0007
i-Butane	lbmole/hr	0.0008	0.0003	0.0000	0.0005
Ethane	lbmole/hr	0.0010	0.0002	0.0006	0.0001
H2O	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	718.5918	240.0000	354.5880	124.0037

Stream		112	113	116	129
Methane	lbmole/hr	0.1183	8.1916	16.2721	24.4637
Nitrogen	lbmole/hr	0.0004	0.0688	0.1337	0.2025
CO2	lbmole/hr	24.7679	90.8482	56.3480	147.1962
SELEXOL	lbmole/hr	0.0000	0.0000	0.0000	0.0000
M-Mercaptan	lbmole/hr	0.0004	0.0001	0.0000	0.0001
Propane	lbmole/hr	0.0001	0.0002	0.0001	0.0004
Refrig-12	lbmole/hr	0.0004	0.0004	0.0000	0.0004
Refrig-40	lbmole/hr	0.0004	0.0002	0.0000	0.0002
VinylCl	lbmole/hr	0.0004	0.0002	0.0000	0.0002
COS	lbmole/hr	0.0002	0.0003	0.0000	0.0003
H2S	lbmole/hr	0.0001	0.0002	0.0001	0.0003
Toluene	lbmole/hr	0.0004	0.0000	0.0000	0.0000
Cl3-C2=	lbmole/hr	0.0004	0.0000	0.0000	0.0000
Refrig-22	lbmole/hr	0.0004	0.0002	0.0000	0.0002
n-Butane	lbmole/hr	0.0004	0.0004	0.0000	0.0004
n-Pentane	lbmole/hr	0.0004	0.0001	0.0000	0.0001
i-Pentane	lbmole/hr	0.0004	0.0002	0.0000	0.0002
22-Mpropane	lbmole/hr	0.0004	0.0003	0.0000	0.0003
i-Butane	lbmole/hr	0.0002	0.0003	0.0000	0.0004
Ethane	lbmole/hr	0.0000	0.0001	0.0002	0.0003
H2O	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	24.8918	99.1119	72.7543	171.8662

Acrion CO2 Wash + SELEXOL to produce CO2 + Pipline Gas from LFG

Stream		130	131	133r	134
Methane	lbmole/hr	21.4863	0.2206	0.0072*	21.2658
Nitrogen	lbmole/hr	0.0617	0.0001	0.0000*	0.0617
CO2	lbmole/hr	94.3190	18.3330	8.2862*	75.9859
SELEXOL	lbmole/hr	210.9525	210.9525	210.9525*	0.0000
M-Mercaptan	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Propane	lbmole/hr	0.0002	0.0000	0.0000*	0.0001
Refrig-12	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Refrig-40	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
VinylCl	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
COS	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
H2S	lbmole/hr	0.0001	0.0000	0.0000*	0.0001
Toluene	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Cl3-C2=	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Refrig-22	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
n-Butane	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
n-Pentane	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
i-Pentane	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
22-Mpropane	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
i-Butane	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Ethane	lbmole/hr	0.0003	0.0000	0.0000*	0.0002
H2O	lbmole/hr	0.0000	0.0000	0.0000*	0.0000
Total:	lbmole/hr	326.8201	229.5063	219.2460*	97.3139

Stream		135	137
Methane	lbmole/hr	21.4792	0.2134
Nitrogen	lbmole/hr	0.0617	0.0001
CO2	lbmole/hr	86.0246	10.0387
SELEXOL	lbmole/hr	0.0000	0.0000
M-Mercaptan	lbmole/hr	0.0000	0.0000
Propane	lbmole/hr	0.0001	0.0000
Refrig-12	lbmole/hr	0.0000	0.0000
Refrig-40	lbmole/hr	0.0000	0.0000
VinylCl	lbmole/hr	0.0000	0.0000
COS	lbmole/hr	0.0000	0.0000
H2S	lbmole/hr	0.0001	0.0000
Toluene	lbmole/hr	0.0000	0.0000
Cl3-C2=	lbmole/hr	0.0000	0.0000
Refrig-22	lbmole/hr	0.0000	0.0000
n-Butane	lbmole/hr	0.0000	0.0000
n-Pentane	lbmole/hr	0.0000	0.0000
i-Pentane	lbmole/hr	0.0000	0.0000
22-Mpropane	lbmole/hr	0.0000	0.0000
i-Butane	lbmole/hr	0.0000	0.0000
Ethane	lbmole/hr	0.0003	0.0000
H2O	lbmole/hr	0.0000	0.0000
Total:	lbmole/hr	107.5661	10.2522

```
 Stream                               100           2           3           4
Vapour frac.                       1.0000      1.0000      1.0000      0.9995
Temperature F                    70.0000*     74.6180    305.4935   100.0000*
Pressure    psia                 14.7000*     14.7000    55.0000*     50.0000
Molar Flow  lbmole/hr           450.7188*    558.2162    558.2162    558.2162
Mass Flow   lb/hr              12833.9648  16963.0686  16963.0686  16963.0686
LiqVol Flow barrel/day          1616.5602   2009.3619   2009.3619   2009.3619
Enthalpy    Btu/hr             1.89100E+06 2.36194E+06 3.56491E+06 2.46739E+06

 Stream                                 5           6           7           8
Vapour frac.                       0.0000      1.0000      1.0000      0.9860
Temperature F                    100.0000    100.0000    336.4234   100.0000*
Pressure    psia                  50.0000     50.0000   185.0000*    180.0000
Molar Flow  lbmole/hr              0.2855    557.9308    557.9308    557.9308
Mass Flow   lb/hr                  5.1435  16957.9253  16957.9253  16957.9253
LiqVol Flow barrel/day             0.3529   2009.0089   2009.0089   2009.0089
Enthalpy    Btu/hr             -4043.8778 2.47143E+06 3.70709E+06 2.27026E+06

 Stream                                 9          10          11          12
Vapour frac.                       0.0000      1.0000      1.0000      0.0000
Temperature F                    100.0000    100.0000    100.0000    100.0000
Pressure    psia                 180.0000    180.0000    180.0000    180.0000
Molar Flow  lbmole/hr              7.8301    550.1007    546.7137      3.3870
Mass Flow   lb/hr                141.1231  16816.8018  16755.7856     61.0165
LiqVol Flow barrel/day             9.6842   1999.3247   1995.1383      4.1864
Enthalpy    Btu/hr           -110828.3538 2.38109E+06 2.36733E+06 -47962.0882

 Stream                               101         102         103         104
Vapour frac.                       1.0000      1.0000      1.0000      1.0000
Temperature F                     96.9662    289.7497   100.0000*     8.0000*
Pressure    psia                 175.0000   515.0000*    510.0000    505.0000
Molar Flow  lbmole/hr            718.5918    718.5918    718.5918    718.5918
Mass Flow   lb/hr              23632.7049  23632.7049  23632.7049  23632.7049
LiqVol Flow barrel/day          2622.8889   2622.8889   2622.8889   2622.8889
Enthalpy    Btu/hr             3.08312E+06 4.31860E+06 2.88974E+06 2.16207E+06

 Stream                               105         106         107         108
Vapour frac.                       0.0000      0.0000      1.0000      1.0000
Temperature F                     -6.7270     -8.0986    -57.0012    -0.0000*
Pressure    psia                 505.0000    500.5263    500.0000    495.0000
Molar Flow  lbmole/hr            124.0037    240.0000    354.5880    354.5880
Mass Flow   lb/hr               5224.0973  10105.1113   8303.4967   8303.4967
LiqVol Flow barrel/day           452.8435    876.4132   1293.6322   1293.6322
Enthalpy    Btu/hr           -260706.9993 -512010.0540 900410.9372 1.12666E+06

 Stream                               109         110         111        111a
Vapour frac.                       1.0000      1.0000      0.2098      0.7993
Temperature F                     12.1267    89.5000*    -45.6218   -34.0000*
Pressure    psia                 490.0000    485.0000   180.0000*   180.0000*
Molar Flow  lbmole/hr            247.0138    247.0138    124.0037    124.0037
Mass Flow   lb/hr               4170.8839   4170.8839   5224.0973   5224.0973
LiqVol Flow barrel/day           900.5503    900.5503    452.8435    452.8435
Enthalpy    Btu/hr            844371.2857 1.02647E+06 -260706.9993 209466.8001
```

Stream		112	113	114	114a
Vapour frac.		0.0000	1.0000	1.0000*	1.0000
Temperature	F	-34.0000	-34.0000	-28.1048	-61.6383
Pressure	psia	180.0000	180.0000	175.0000	25.0000*
Molar Flow	lbmole/hr	24.8918	99.1119	24.8918	24.8918
Mass Flow	lb/hr	1092.3377	4131.7600	1092.3377	1092.3377
LiqVol Flow	barrel/day	90.8994	361.9441	90.8994	90.8994
Enthalpy	Btu/hr	-80078.1141	289544.9030	73587.5945	73587.5945

Stream		115	116	117	118
Vapour frac.		1.0000	1.0000	0.0000	0.1330
Temperature	F	89.5000*	-42.9899	-29.1257	-67.0000*
Pressure	psia	20.0000	180.0000	180.0000	78.2974
Molar Flow	lbmole/hr	24.8918	72.7543	167.2458	167.2458
Mass Flow	lb/hr	1092.3377	2744.6745	7360.4366	7360.4366
LiqVol Flow	barrel/day	90.8994	265.7624	610.6508	610.6508
Enthalpy	Btu/hr	107589.1974	210664.9313	-523303.8934	-523303.8934

Stream		120	121	122	123
Vapour frac.		0.0000*	0.1645	1.0000*	1.0000
Temperature	F	-1.4316	-67.0000	-66.9997	50.0000*
Pressure	psia	300.0000*	78.2974	78.2974	73.2974
Molar Flow	lbmole/hr	71.9710*	239.2168	239.2168	239.2168
Mass Flow	lb/hr	3167.4203	10527.8563	10527.8563	10527.8563
LiqVol Flow	barrel/day	262.7810	873.4318	873.4318	873.4318
Enthalpy	Btu/hr	-173715.8792	-697019.7870	667847.2050	929936.6832

Stream		124	125	126	127
Vapour frac.		1.0000	1.0000	1.0000*	0.0000*
Temperature	F	296.4501	100.0000*	-0.4682	-1.4325
Pressure	psia	315.0000*	310.0000	305.0000	300.0000
Molar Flow	lbmole/hr	239.2168	239.2168	239.2168	239.2168
Mass Flow	lb/hr	10527.8563	10527.8563	10527.8563	10527.8563
LiqVol Flow	barrel/day	873.4318	873.4318	873.4318	873.4318
Enthalpy	Btu/hr	1.46318E+06	970666.3027	708570.2459	-577479.2062

Stream		128	129	129r	130
Vapour frac.		0.9998	1.0000	1.0000	0.0000
Temperature	F	-37.6962	89.5000*	89.5000*	32.6953
Pressure	psia	180.0000	175.0000	175.0000*	495.0000
Molar Flow	lbmole/hr	171.8662	171.8662	171.8781*	326.8201
Mass Flow	lb/hr	6876.4343	6876.4343	6876.9192	61454.5813
LiqVol Flow	barrel/day	627.7066	627.7066	627.7506	4206.2638
Enthalpy	Btu/hr	500209.8301	715742.0525	715791.6453	-8.19219E+06

Stream		130a	131	131a	132
Vapour frac.		0.2978	0.0000	0.0447	1.0000
Temperature	F	10.4936	10.4936	7.1531	10.4936
Pressure	psia	20.0000*	20.0000	7.0000*	20.0000
Molar Flow	lbmole/hr	326.8201	229.5063	229.5063	97.3139
Mass Flow	lb/hr	61454.5813	57767.5558	57767.5558	3687.0277
LiqVol Flow	barrel/day	4206.2638	3850.6484	3850.6484	355.6156
Enthalpy	Btu/hr	-8.19219E+06	-8.54566E+06	-8.54566E+06	353500.5976

```
 Stream                              133          133a          133r           134
Vapour frac.                       0.0000        0.0000        0.0000        1.0000
Temperature F                      9.8431        7.1531       9.8928*      89.5000*
Pressure     psia               490.0000*        7.0000     490.0000*       15.0000
Molar Flow   lbmole/hr           219.2541      219.2541     219.2460*       97.3139
Mass Flow    lb/hr             57322.3256    57322.3256    57321.9682     3687.0277
LiqVol Flow  barrel/day         3813.2116     3813.2116     3813.1819      355.6156
Enthalpy     Btu/hr          -8.47568E+06  -8.58252E+06  -8.47454E+06   422938.0312

 Stream                              135          135R           136           137
Vapour frac.                       1.0000        1.0000        1.0000        1.0000
Temperature F                     93.4851      93.4866*        7.1531      130.6391
Pressure     psia                 15.0000      15.0000*        7.0000      16.0000*
Molar Flow   lbmole/hr           107.5661     107.4974*       10.2522       10.2522
Mass Flow    lb/hr              4132.2549     4129.1038      445.2270      445.2270
LiqVol Flow  barrel/day          393.0523      392.8017       37.4366       37.4366
Enthalpy     Btu/hr           471239.1389   470941.4698    36851.8182    48301.1218

 Stream                            Wcomp1        Wcomp2        Wcomp3         wpump
Enthalpy     hp                  472.7845      485.6331      485.5625       41.9914

 Stream                          wvaccomp      Wco2comp
Enthalpy     hp                    4.4997      209.5752

 Stream                             qair1         Qair2         qair3       Qairco2
Enthalpy     Btu/hr           1.09752E+06   1.43682E+06   1.42886E+06   492520.5527

 Stream                              QHX1         QCOND          QREB          QHX2
Enthalpy     Btu/hr           727674.5180   2.03442E+06   199380.2158   226256.7636

 Stream                              QHX3          QHX4          QHX5          QHX6
Enthalpy     Btu/hr           182099.4256   153665.7035    34001.6021   1.36486E+06

 Stream                              QHx7          QHX8          QHX9         QHX10
Enthalpy     Btu/hr           262089.4702   262096.0457   1.28604E+06   215532.2154

 Stream                             QHX11         qdehi         Qhx12
Enthalpy     Btu/hr            69437.4363    61722.6696   470173.7848

 Stream                              QHX4          QHX9
Enthalpy     tons*                12.8055      107.1707

 Stream                        Refrig-20
Enthalpy     tons*                94.3652
```

Figure A.1

LFG to Pipeline Gas and Carbon Dioxide

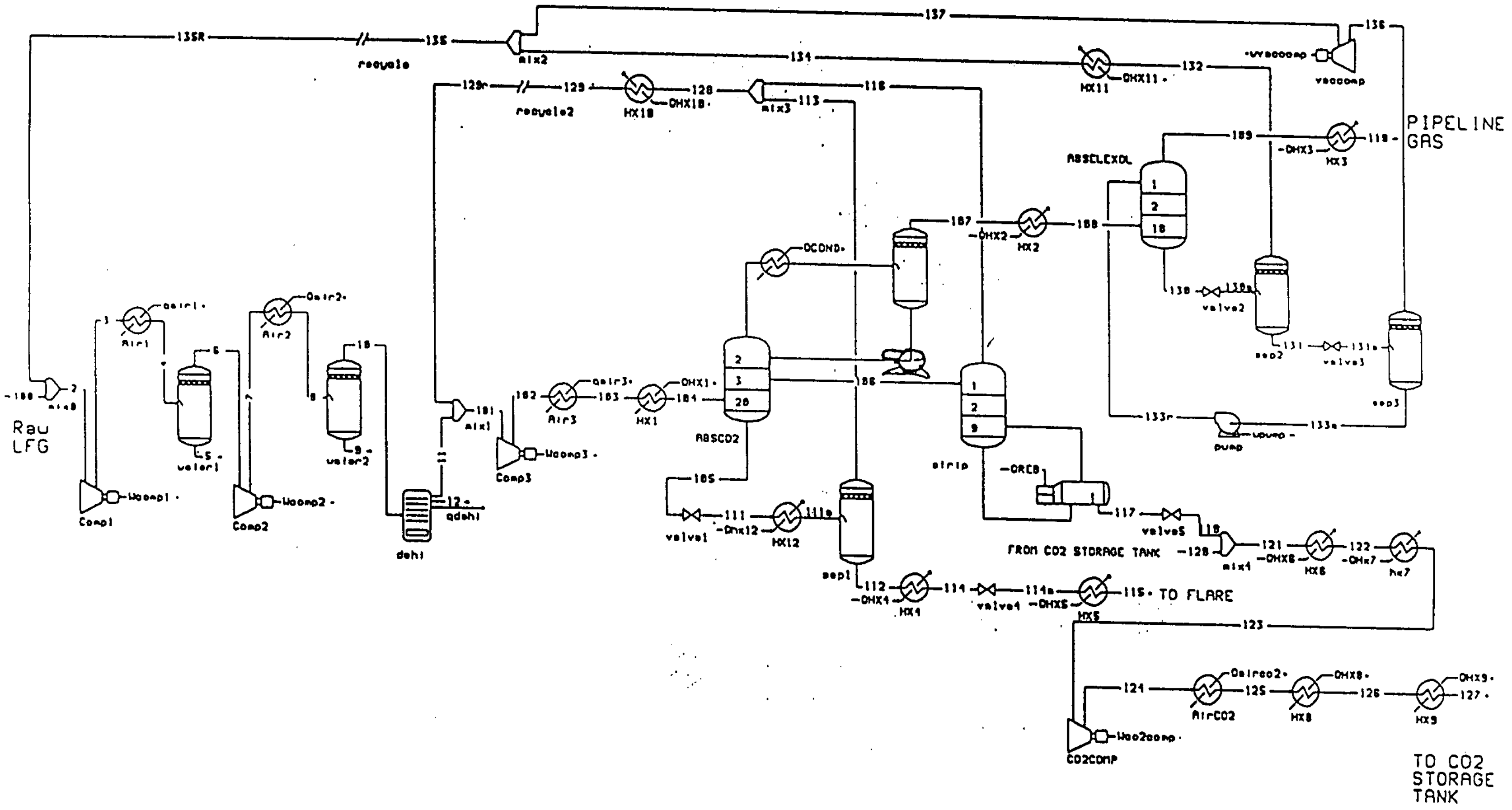
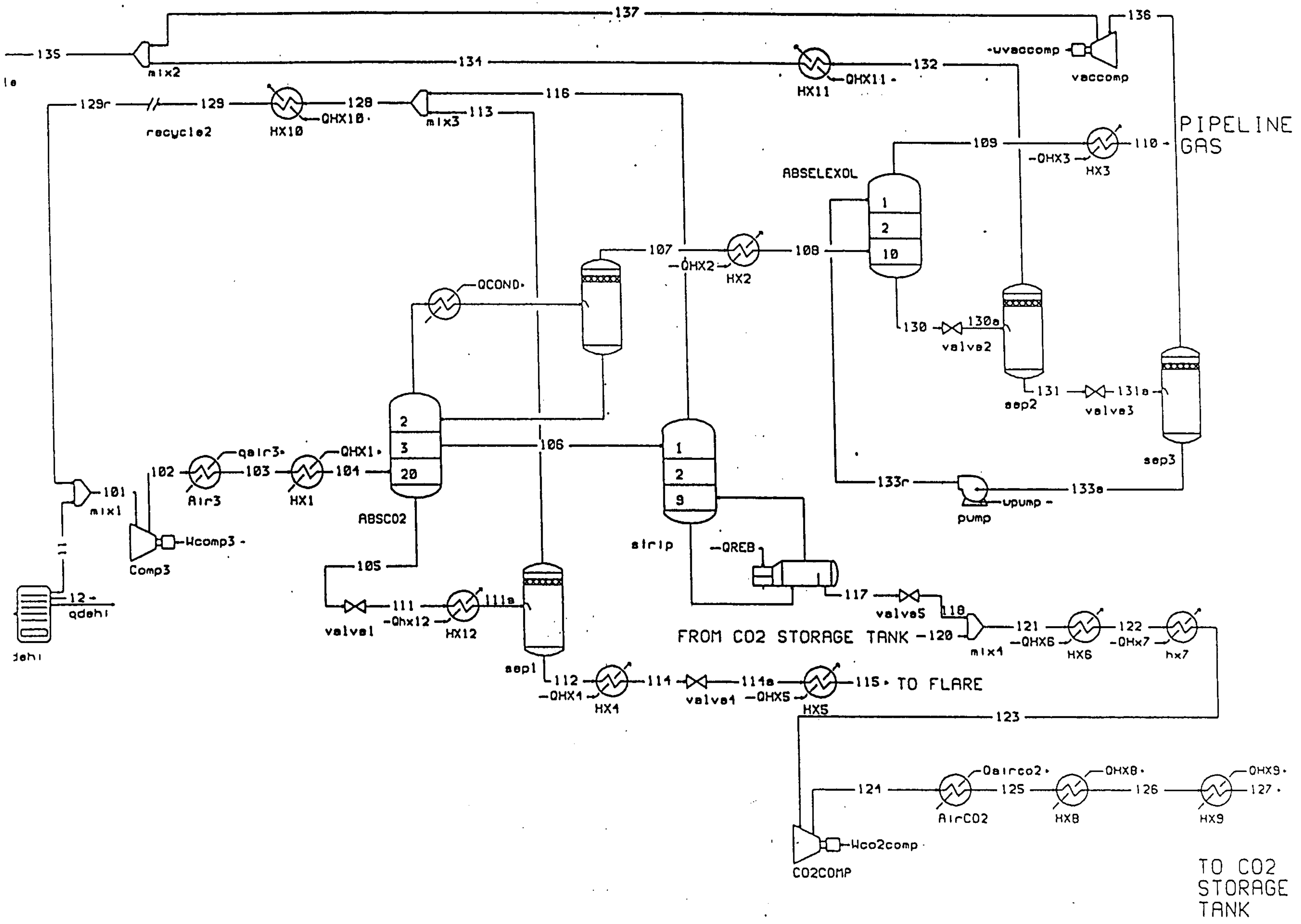


Figure A.2

Detail of CO2 Wash and Selexol



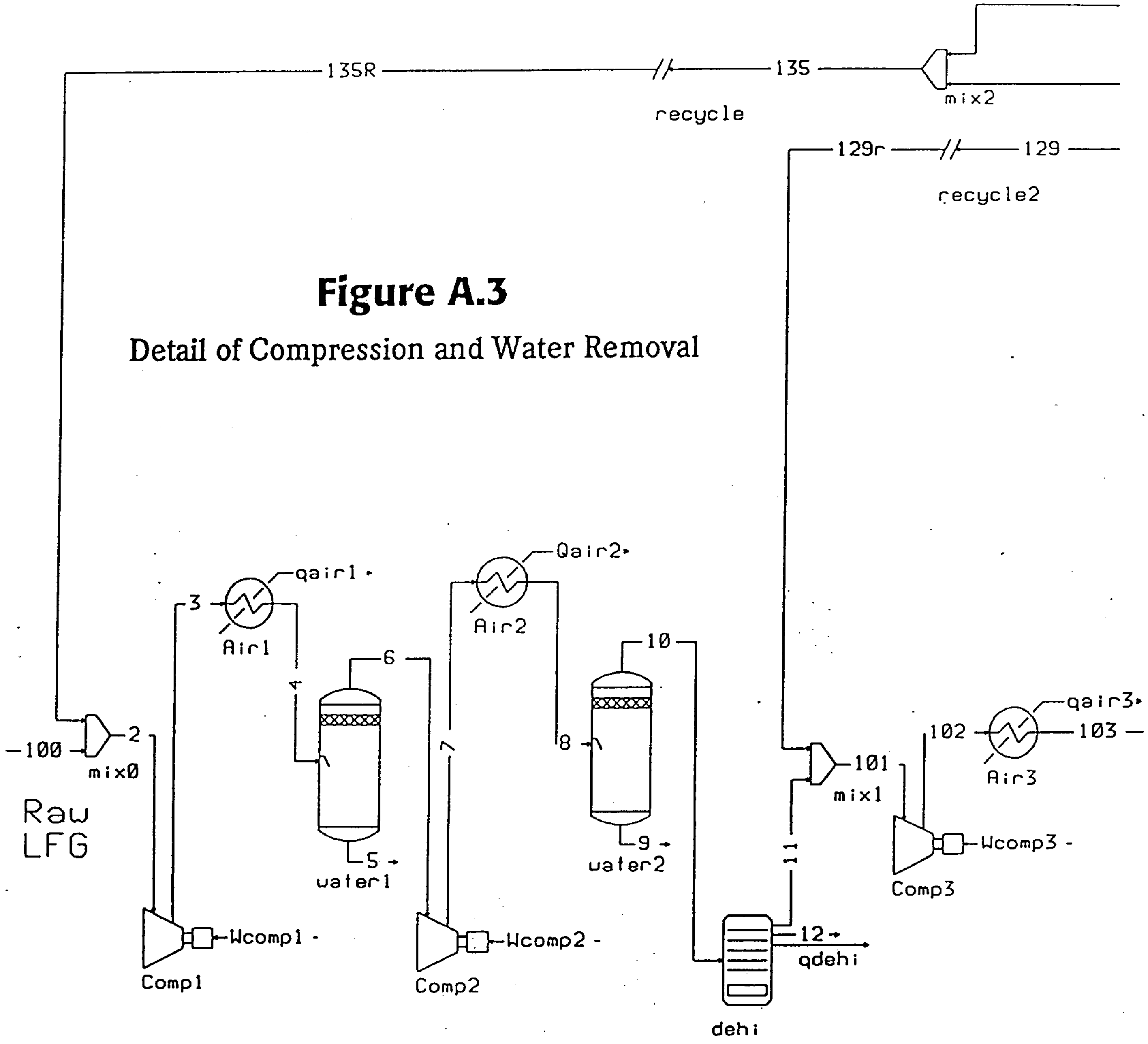


Figure A.3

Detail of Compression and Water Removal

Figure A.4

Detail of HX1 Feed Cooler

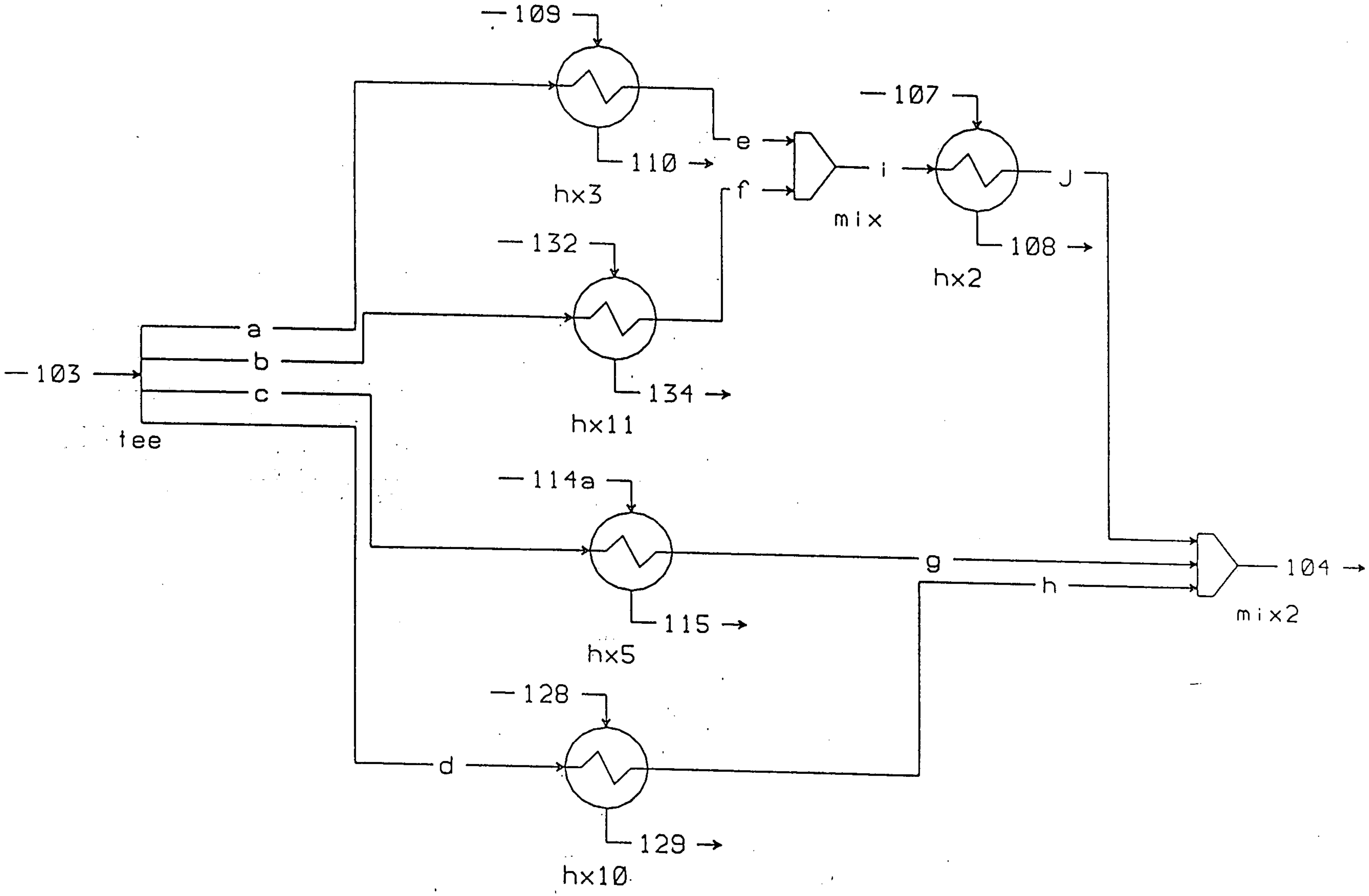
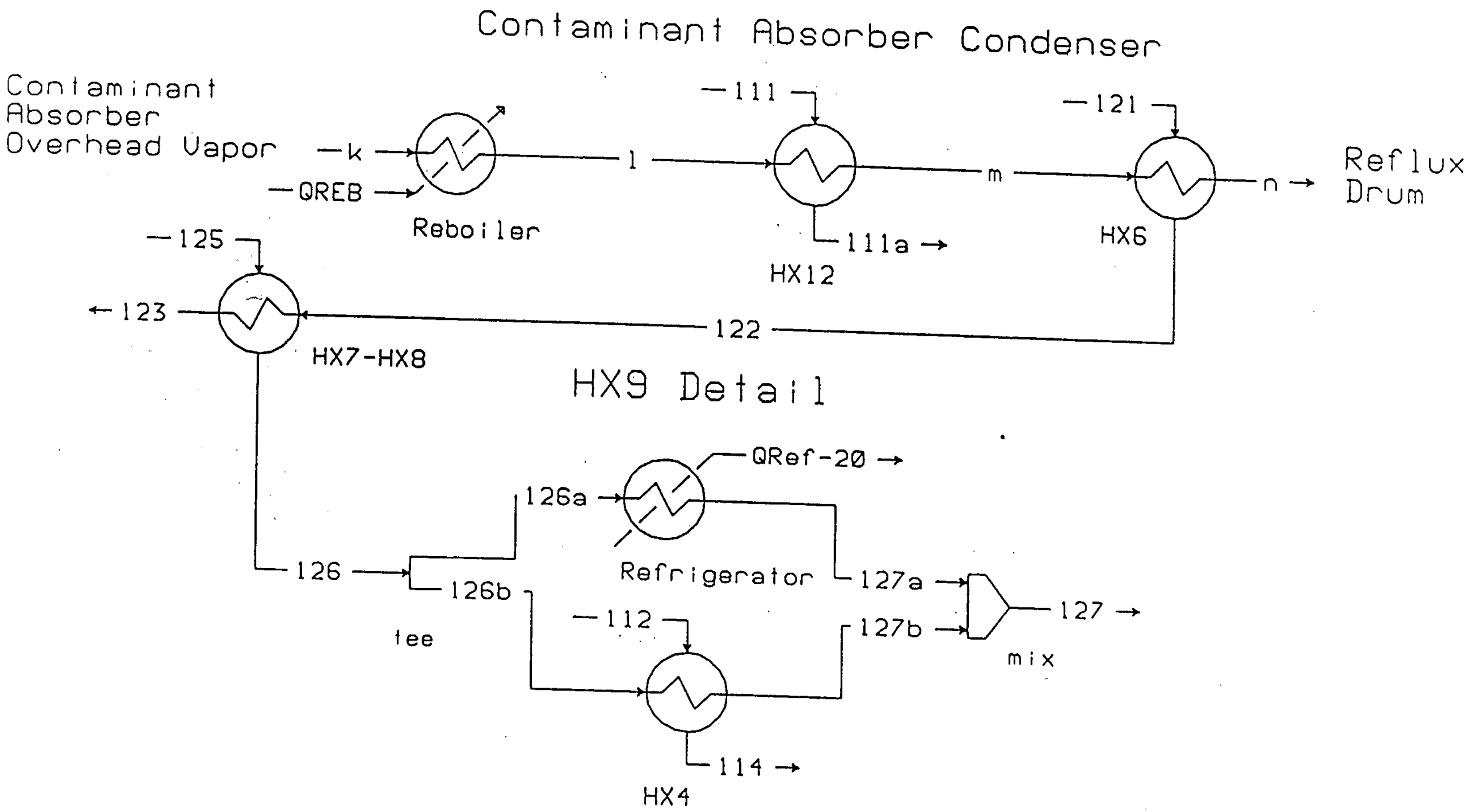


Figure A.5

Heat Exchanger Details



APPENDIX B

Final Report
Swapping LFG Methane for Remote LNG

by
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April 4, 2000

Mr. William R. Brown
President
Acrion Technologies, Inc.
9099 Bank Street
Cleveland, Ohio 44125

Re: LFG-to-"NG"-to-LNG

Dear Bill:

This letter constitutes a Final Report by Vandor + Vandor, per Subtasks 2.1.2 and 2.1.3 of our subcontract with Acrion Technologies, Inc., concerning Acrion's study of Landfill Gas to LNG and LCO_2.

Background

The Vandor + Vandor subcontract with Acrion Technologies, Inc., is based on several core assumptions, including the following:

- Many landfill sites, including publicly owned ones in Maryland, flare landfill gas (LFG). While less harmful than allowing methane and carbon dioxide to escape directly into the atmosphere, flaring wastes resources, and is not emission free.
- The use of Liquid Natural Gas (LNG) as a vehicle fuel, especially by heavy duty trucks and buses, can reduce vehicular emissions and the need for imported fuel.
- While the direct production of LNG from LFG may not yet be economically viable, opportunities exist for linking LFG clean-up systems, (such as that developed by Acrion Technologies), with the production of LNG at underutilized "peak shaving" plants.
- The trading ("wheeling") of cleaned LFG, in exchange for off-site produced LNG, would advance pollution prevention at both the landfill and within the area served by the LNG fleet.
- By capturing the inherent value of the LFG methane stream, a municipality can create an economically viable "loop" that links its landfills to its fleets.

Vandor + Vandor's subcontract focused on Maryland and the following topics as a way of examining the above outlined assumptions:

- Pipeline company standards for accepting cleaned-LFG.
- BGE's gas distribution system.
- Peak shaving liquefaction / LNG distribution and availability.
- LFG-to-pipeline-gas, traded for LNG?
- LFG-to-electricity, traded for LNG?
- LFG-to-pipeline-gas, or LFG-to-electricity?

Subtask 2.1.2 -- Pipeline company gas requirements.
The attached letter from Baltimore Gas and Electric (BGE), dated December 2, 1999, responds to our request for information regarding BGE's requirements for "pipeline quality" gas. The following is a summary of those requirements:

1. **The upper limit for carbon dioxide and oxygen:** None, other than the BTU content. They do not test for CO_2 nor oxygen, only for nitrogen.
2. **Minimum BTU content of the gas:** 1000 BTUs. The actual average quality of BGE's gas runs from 1034 to 1042 BTUs. Please find attached to this report a copy of "Gas Analysis -- Component Average Report," which tabulates, by month, 10 chemical components of pipeline gas at three gate stations.
3. **Wobbe index:** They do not use a Wobbe index.
4. **Upper limit for hydrogen sulfide, and total sulfur:** They "odorize" the gas with sulfur compounds up to 3 parts per million. That limit will "allow the gas to burn without leaving an odor."
5. **Typical pipeline pressure:** BGE's "high pressure" lines range from 70 psig to 100 psig. Their "over high pressure" lines range between 100 psig and 300 psig.
6. **Estimated cost of connecting to a gas main:** $800 to $9,500.
7. **Estimated cost of building a pipeline:** $8 to $71 per foot of new pipeline, depending on pipe size and field conditions.
8. **Range of values for purchased gas:** Varies with the market and seasons, but has fluctuated between $1.40 to $4.00 per million BTUs between 1993 and 1998. Assume that cleaned LFG will not sell for more than $2.00 per million BTUs.

BGE's gas distribution system.
Attached are four copies of the BGE's Gas Distribution System in Maryland. These maps will allow the list of Maryland landfills (previously sent to Acrion) to be matched to the pipeline system. In addition, the maps show the "gaps" in the distribution system. Those areas within BGE's "gas franchise area" not served by pipelines are possible "opportunity" zones for LNG sales, including for vehicle use.

For example, if public policies encouraged AFVs in Union Bridge, New Windsor or Taneytown, (all in the north-west corner of BGE's franchise area), LNG and/or LNG-to-CNG could be practical if LNG could be delivered, stored and distributed at a cost that is competitive with diesel, gasoline, and propane. By contrast, pipeline-fed CNG would not be an option because of the absence of a pipeline network.

Subtask 2.1.3 -- Peak shaving liquefaction / LNG distribution.
Item 9 in the 12/2/99 BGE letter establishes a sales price of 33¢ per gallon of LNG. In conversations with BGE staff (including Tom Wieczynski, at [410] 597 6406), we can assume that the 33¢ price could change (say to 36¢) as pipeline gas prices fluctuate,

and that it does not include the cost of transporting the LNG from BGE's Baltimore LNG plant to the customer. We should assume that the "delivered" price of LNG, within 100 miles of Baltimore, will be approximately 42¢, where 5¢ to 7¢ of that amount is the cost of transporting LNG by truck.

The conservative, untaxed cost of LNG as a vehicle fuel (say for a public fleet) would be approximately 84¢ per "diesel equivalent" gallon, assuming that a spark-ignited heavy duty LNG engine will use approximately two gallons of LNG to move a bus/truck the same distance as does one gallon of diesel in a standard vehicle.

A more "optimistic" cost of LNG (perhaps when used by "dual fuel" diesel-pilot-ignition engines), might be 71.4¢ per diesel equivalent gallon, where the "alternative fuel vehicle's" (AFV) engine uses 1.7 gallons of LNG to move it the same distance as one gallon of diesel moves its counterpart.

Both the 84¢ and 71.4¢ figures may be "competitive" with diesel, especially if public policy considerations regarding air quality and domestic fuel use are part of the decision-making process. On the other hand, the notion that rising diesel prices may increase the gap between LNG and diesel is not supported by historic oil and natural gas price trends. The two commodities seem to rise and fall in general unison.

BGE's most efficient LNG plant uses a "turbo expander" for refrigeration power. The low operating cost of such plants allows BGE to charge a relatively low price for "excess" LNG produced and sold beyond its own need. However, BGE staff has expressed several constraints on their ability to regularly sell LNG. For example, because the primary function of their LNG plant is for "peak shaving", they cannot commit to a steady delivery cycle, but only to sales on a "best effort" basis. Similarly, their LNG "surplus" will vary by time of year and will depend on their peak shaving needs.

Still, BGE will confirm that they routinely have excess LNG production capacity, beyond their peak shaving needs and beyond their ability to store product, despite the fact that the plant does not run 365 days per year. Indeed, conversations with BGE staff confirm that if an LNG market developed they would examine the possibilities for enhancing the efficiency of their LNG production cycle, increasing production without adding equipment or staffing costs.

One constraint on BGE's turbo expander LNG plant is the lack of demand during the summer for the low-pressure "outflow" gas in the pressure letdown refrigeration process. Vandor + Vandor has sketched out a solution to this problem and has had informal conversations with BGE staff regarding their interest in fully utilizing the plant's LNG production capacity. Obviously, even if our approach proved viable, BGE's interest would depend on a steady market for LNG, beyond their peak shaving need.

We can safely predict that if a customer sought a steady supply of LNG from BGE, such as for a fleet, a dependable program of deliveries could be established, even within BGE's current LNG production cycle.

However, in order to insure that the customer's LNG storage tanks would never be fully empty, it might need to have a greater LNG storage capacity (on a diesel gallon equivalent basis) than a similar diesel dispensing facility. In addition, the customer would be prudent to establish a second LNG source, such as Philadelphia Gas Works, so that a consistent delivery schedule and competitive pricing can be maintained.

Those two suggestions -- a large enough LNG storage facility that acts as "valve" between different rates of LNG input and output, and is served by multiple LNG suppliers -- is the subject of a Vandor + Vandor response to a recently issued New York State Energy Development Authority (NYSRDA) Program Opportunity Notice (PON), seeking technical proposals for innovative gas storage models. That study will begin in late-January, 2000, and will focus on gas storage and distribution issues in New York State. We are confident that the findings of that study will be applicable to Maryland, Pennsylvania and other north-east states.

The viability of a regional LNG storage/distribution center would be enhanced if large heavy-duty fleets became steady customers for the dispensed LNG. In turn, the ability of a public bus fleet, for example, to fund an LNG-fueled AFV program may be enhanced if LFG-flaring at landfills is replaced by LFG-cleaning and "wheeling."

A comprehensive plan that links landfill sites to AFVs may include several steps (clean-up of the LFG, wheeling it for LNG, storing and distributing LNG, dispensing it to AFVs...), each of which can function independently and at an appropriate scale. For example, if at a given time only five or six LNG trash haulers require product, the distribution of LNG from BGE's plants can still move forward, so long as the dispensing system that fuels the trucks has adequate storage capacity and a second source of LNG. The wheeling of LFG-produced energy can occur at its optimum rate, independently of the quantity of dispensed LNG.

LFG-to-pipeline-gas, traded for LNG?

This study assumes that the value of recycled LFG can be greater than merely flaring it, even though flaring is relatively inexpensive when compared to cleaning LFG to pipeline quality and/or using it on-site to produce electricity.

A core aim of this study was to establish the extent to which Acrion's clean-up system might economically convert LFG to pipeline quality gas, which in turn could be traded for LNG. Simply stated, the issues boil down to the following:

- Can the Acrion process cost-effectively produce pipeline quality gas (accounting for all capital, financing, operating and LFG costs), even though that gas will not likely have a value of more than $2 per million BTUs?
- If so, is the best "trade" achieved by exchanging a million BTUs of pipeline quality gas for approximately 5.555 gallons of LNG, where we assume that LNG would be valued at 36¢ per gallon?
- Can Acrion's CO_2 production capabilities enhance this model in the current market for CO_2?
- Can the evolving market for emission reductions generate enough "value" to enhance the model?

LFG-to-electricity, traded for LNG?

An alternative approach would use LFG as a fuel to produce electricity, and trade "kilo-watt-hours" for LNG. A modified, and possibly less costly, Acrion clean-up process might yield benefits related to the operating life of the electric generating equipment and might yield more emission trading "credits" than flaring or standard LFG-to-electricity methods, but at lower capital, financing, and operating costs than the LFG-to-pipeline-gas model. The absence of CO_2 production in a modified Acrion process may not be a debit if there is no existing market for CO_2.

In evaluating the capital, financing, operating costs and product "values" associated with trading electricity for LNG, the following assumptions might be considered:

- The value of the LFG should be no more than 25¢ per million BTU, and in the context of municipal landfills that now flare LFG, the value should approach 0¢;
- LNG can be purchased from BGE for say, 36¢ per gallon, untaxed;
- BGE will not likely "value" electricity delivered to it at more than say, 3¢ per KWH;
- the capital costs for an LFG-clean-up project should be amortized in less than 10 years for public sites, and less than 5 years for private market sites; and
- we can assume a market interest rate on capital of say, 11%.

Several questions follow:

- Is it best to wheel 12 KWHs of electricity for one gallon of LNG, where 12 times 3¢ per KWH buys one gallon of 36¢-LNG, or to sell/use the electricity directly and purchase the LNG outright?
- Can the evolving market for emission reduction generate enough additional "value" to enhance the LFG-to-electricity model, and if so, is it a lesser or greater value than that achieved by Acrion's LFG-to-NG clean-up?

- Can a public entity establish a higher value for LFG-to-electricity than 3¢ per KWH, perhaps by establishing an "avoided cost of power" and wheeling the electricity directly to its own use, even if it must pay "transportation" or "interconnection" or "demand charges" to the local utility?

LFG-to-pipeline-gas, or LFG-to-electricity?

In comparing the two models, we can assume that the capital and operating cost of bringing LFG to pipeline-quality will likely be **more** than merely cleaning the LFG to optimum turbine or engine standards. Conversely, electricity generation will likely require less capital and financing costs over a 5 or 10 year amortization period.

Two additional questions need to be examined:

1. Can Acrion add value to the production of 3¢ per KWH electricity, by reducing emissions and/or by increasing the efficiency and longevity of the electric generating equipment; or
2. should Acrion produce pipeline quality gas, via a more expensive system, and trade a "gallon equivalent" of cleaned-LFG (57,000 BTUs), valued as "pipeline gas" at 11.4¢ (at $2.00 per million BTU), for a gallon of LNG priced at 36¢?

Conclusions

As a public policy matter, the trading of LFG-produced electricity or pipeline gas for LNG is especially warranted when the LNG is used in AFVs. The benefit of the LFG-to-LNG model -- either with pipeline quality gas or with electricity -- is that both landfill emissions **and** vehicle emissions are reduced. LFG is a "renewable" and domestic power source.

If there is a market for emission reductions, the LFG-to-NG model should yield higher values. However, if there is no market for liquid CO2, the economics of Acrion's full clean-up system may not be favorable.

In the event that direct wheeling of KWH for LNG is the most cost-effective option, Acrion needs to demonstrate that its process can offer tangible benefits to the operator of an LFG-fed turbine, such as lower operating costs, longer equipment life, and a cost effective way to capture greater emission reduction values.

Sincerely,

David Vandor

APPENDIX C

Final Report
Characterization of Sewage Digester Gas (Biogas)

by
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BNL Phase II

prepared for

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April 2000

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BNL Phase II

A. SCOPE of WORK

A.1. Chapter 1: Digester Gas (**Biogas**) as a Renewable Source of Energy

Acrion is interested in digester gas as a renewable source of methane for direct or indirect production of liquid methane.

AplusB will provide the following:

1.1 an introduction to waste water treatment (WWT) plants

1.2 an investigation of the characteristics of sewage digester gas

1.3 a working data base of commercial, or near commercial, technologies and companies associated with bio digesters,

1.4 an investigation of how sewage digester gas is currently managed and utilized in a sample of the USA territory. Existing or planned demonstration plants will be sought,

1.5 an investigation of the use of CO_2 for sewage effluent pH adjustment

1.6 addendum

A.2. Chapter 2: Western Europe & Bio-Digesters

Acrion is interested in the use of, and advances in, anaerobic digestion systems as a source of biogas, and its treatment, for productive use in Western Europe.

AplusB will:

2.1 Provide an introduction to waste water treatment (WWT) plants in Europe

2.2 Investigate characteristics of sewage digester gas (biogas)

2.3 Investigate how sewage digester gas is currently managed and utilized in Europe

* * •

B. METHODOLOGY

AplusB will contact one or more "large" WWT (waste water treatment) plants in Ohio and other states to determine:

a) how digester gas is currently managed and utilized,

b) other needs within the sewage treatment process - - e.g. the use of $\underline{CO_2}$ for sewage effluent pH adjustment ,

c) chemical and material balance of sewage plant,

d) factors that influence quantity of digester gas produced and conditions under which these factors may be adjusted to increase flow of digester gas.

* * *

C. REPORT

C. 1. Executive Summary

Anaerobic digestion of organic matter can and does produce "biogas".

Biogas can be produced "naturally", as in landfills, or "artificially" as in digesters. Digesters are found both in some municipal WWT Plants and in some industrial WWT Plants (where they process animal manure, vegetables and fruits, wineries, etc).

Biogas produced by digesters typically includes 65% methane, 35% carbon dioxide, and traces of H_2S (in concentrations generally less than 1%) generated from the conversion of the sulfur compounds within the treated substrate. Trace amounts of nitrogen, hydrogen, methyl mercaptans and oxygen are also typically found in digester biogas.

Landfill biogas (LFG) can include a significant number of other gases, including CFC's and HFC's.

Digesters in Europe operate under a range of technologies such as Valorga (France), BAT (Germany), Dranco (Belgium). In the USA, large Architectural/Engineering firms such as Camp, Dresser and McKee, Black and Veatch have developed specific digester technologies that are installed, for instance, at the recent City of Boston water reclamation plants.

Several large USA metropolitan WWT plants (Boston, Chicago, East Bay Municipal Utility District, etc), operate digesters to stabilize sludge. This process generates volumes of biogas which, in several cases, exceed 2.5 MMSCFD (threshold at which Acrion's technology produces commercially viable products using biogas as feed stock).

Biogas produced by digesters installed in municipal WWT plants is customarily used as a source of in-plant energy to generate electricity, or as a source of heat (to regulate the anaerobic digestive process)

No occurrence of methane/carbon dioxide separation at *municipal* WWT plants was discovered.

Some occurrences of water and H_2S removal from biogas prior to combustion in IC engines or turbines were found.

Few municipal WWT plants which use digesters systematically keep records of the volumes and characteristics of the biogas they produce.

Economic feasibility of digester projects differ in USA and Europe (where environmental preservation requirements, coupled with high energy costs and state grants facilitate cooperative digester projects)

Some *industrial* (not municipal) WWT plants purchase carbon dioxide for in plant use.

Examples of WWT Plants Features

Muni WWT Plants	Boston	Chicago	East Bay	Columbus	Cincinnati	West Point King County	Valenton France
Daily WW volume treated in GPD	1.3 billion	781 million	-	--	-		
Anaerobic Digestion in use	Yes	Yes	Yes	Yes	Yes will be removed in future	Yes	Yes
Daily volume of biogas in MMSCFD	4.08	4.76	unknown	0.800 in 99 1.6 in future	0.6 to 0.8	2.6	1.300
Characteristics of biogas	65% CH_4[1] 33 % CO_2 traces	60 to 65% CH_4 30 to 35% CO_2 traces	63 to 64% CH_4[1] 34% CO_2 traces		60 CH_4[1] 38% CO_2 traces	64 CH_4[1] 32% CO_2 traces	65 CH_4[1] 33% CO_2 traces
In-plant use of biogas	90% for electricity	95% for electricity	100% for electricity	heat for in-plant use	heat for in-plant use	energy recovery	
	4 MW steam turbines		100% of the biogas is for co-generation				
H_2S in biogas	1,000 ppm and higher; reduced to 100 ppm by ferrous chloride	200 to 500 ppm reduced to 10 ppm by Sulfa Treat	unknown	unknown	2 000 ppm	150to 200 ppm	
In-plant CO_2 demand	none	none	none	none	none		
pH adjustment	sodium bicarbonate				ferrous chloride		

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[1] These are estimates – not actual measurements

Conclusions

- Digester technologies, which produce biogas while reducing and stabilizing WWT plant sludge, is gaining favor both in the US and in Europe.
- Characteristics of WWT biogas produced in municipal plants are generally: 65% CH_4, 35% CO_2 and traces of H_2S (30 to 2,000 ppm); WWT plants generally do not record other trace contaminants.
- Municipal WWT plants have resolved the technical and operational aspects of treating waste water under regulatory compliance; these plants now focus on improving their operational economics; many are selling stabilized sludge as a by-product of their process and are avoiding out of pocket expenses by recovering energy from the biogas they generate.
- Although this aspect of WWT plant operations was not investigated, it is logical to assume that those WWT plants which are selling by-products "outside the fence", have organized a marketing/sales function. It would seem probable that a new product line, such as liquid CO_2, if separated out of biogas by the Acrion technology, would be of interest to the larger WWT plants for which they would generate revenue. Informal discussions with a world renowned industrial gas producer showed that there are some geographical areas in the USA where "new" sources of CO_2, if strategically located, would be of interest.
- Unverified estimates from the WWT equipment industry suggest that there might be as many as 250 municipal WWT plants in the USA which generate in excess of 2.5 MMSCFD[2] through anaerobic digestion. This number appears plausible, considering the number of metropolitan areas in the USA with population in excess of 500,000.
- Details of municipal WWT facilities in operation in the Cities of New York, Boston, Chicago and the East Bay Municipalities[3] are included in this report.
- CO_2 from biogas emanating from WWT plants located where CO_2 sources are distant, thus requiring CO_2 to be trucked in, may serve local CO_2 requirements competitively.

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[2] MMSCFD = million standard cubic feet per day

[3] East Bay Municipal Utility District servicing portions of Alameda and Contra Costa in California

C. 2. Report

Chapter 1: Digester Gas as a Renewable Source of Energy

1.1 Introduction: waste water treatment (WWT) plants

Waste water treatment in the USA

In the USA, the volume of domestic waste water produced per capita ranges from 100 to 175 gallons per day, depending on location and season.
Domestic waste water consists of organic and inorganic wastes. Organic wastes can be digested by bacteria and other microorganisms. Inorganic wastes are substances from minerals such as sand, salt, iron, calcium, etc.; the latter are not digestible by microorganisms.

Sludge results from microorganisms having digested the organic portion of waste water; this is an exothermic reaction.
Sludge needs to be stabilized to render it inactive and thus appropriate to be used for commercial applications such as compost. This stabilization process can be performed by anaerobic digestion (AD) which produces biogas, a mixture of methane, carbon dioxide, and other constituents which include nitrogen, oxygen and hydrogen sulfide.

Design flows of typical WWT plants in the USA

WWT plants design flows range from 1.3 million gallons a day (MGD) (the City of Willits, CA) to 415 MGD (East Bay Municipal Utility District, CA.)
Greater design flow rates are found in large metropolitan areas such as the Metropolitan Boston Deer Island facility: 1.3 billion GD. This facility includes state of the art digesters with a capacity of treating 50 tons of sludge per day.

Anaerobic digestion

In 1986, there were 15,400 municipal sewage treatment plants in the USA with a combined flow of 36.6 billion gallons per day [15, page 2-19].

It is estimated, by a major US Vendor of WWT equipment, that 25% of these WWT plants utilize anaerobic digestion technology to process the "sludge" produced.

It is further estimated that, within that 25%, 100 to 200 plants produce biogas in volumes greater than 2.5 MMCFD (million standard cubic feet per day).
This volume is the threshold where investment in the Acrion technology becomes economically viable.
Municipal anaerobic sludge digesters have been recovering the biogas produced for internal use, for years.

Note: there has been an increased interest in controlled anaerobic digestion of municipal solid waste (MSW) as an alternative to landfills and incineration. In 1993, there were 15 plants in operation worldwide with 1 installation in the USA.

The treatment of municipal waste water results in the formation of slurries, high in suspended solids. These slurries are commonly referred as to "sludge". Sludge, an odorous, watery mixture; it is the result of primary and secondary treatment that have removed solids, organic matter and bacteria out of waste water; sludge needs to be stabilized.

Anaerobic digestion of sludge includes two basic steps:

a) the first step is to convert the organic material to volatile acids,

b) the second step is to convert the volatile acids to methane. The bacteria that convert the organic materials into volatile acids are called acid formers. Since acids are being formed, pH becomes very important. The anaerobic process requires that the pH stays as close to 7.0 as possible. The volatile acids are changed into methane by a second set of bacteria (methanogenic bacteria).

Most digesters operate at 95°F or 35°C using mesophilic bacteria.

The biodegradable portion of the organic fraction of MSW can be converted biologically under anaerobic conditions to a gas containing carbon dioxide and methane (CH_4). The principal end products are carbon dioxide, methane, ammonia, hydrogen sulfide, and resistant organic matter. In most anaerobic conversion processes carbon dioxide and methane constitute over 99% of the total gas produced. The resistant organic matter (or digested sludge) must be dewatered before it can be disposed of by land spreading or landfilling. Dewatered sludge is often composted aerobically to stabilize it further before application [1].

Anaerobic digestion, technologies for the production of methane and humus product from organic materials

In recent years, great interest, especially in Europe, has been shown for the processing of organic waste because of the opportunity to recover methane and to produce compost. Summaries of anaerobic digestion processes and technologies are included in Table A, enclosed.

Bio-digesters, environmental factors

To maintain an anaerobic treatment system that will stabilize organic waste efficiently, the bacteria must be in a state of dynamic equilibrium; the reactor contents should be free of undesirable constituents and the pH of the aqueous environment should range from 6.5 to 7.5. The pH should not drop below 6.2 because the methane bacteria cannot function below this point. Other conditions involving nutrients and temperature must also occur [1, p. 681].

Biogas usage

Some larger plants use sludge-produced biogas to power gas engines.
In small plants, disposal of sludge gas is normally by burning.
Other methods of gas disposal exist [2, p 193]

Engines, biogas cleanup

Opinions as to the necessity of "cleaning " biogas before it is utilized to fuel internal combustion engines differ considerably.

There is, however, consensus that biogas H2S must be removed as well as water.

Beyond these two obligations, other biogas contaminants removal processes are implemented within a wide range.

The author of this report witnessed, in the field, extensive damage (resulting in total failure) to engine turbo chargers from silicates originating from biogas from which contaminants had not been removed.

Engine manufacturers are concerned with contaminants, including water and H2S that may be included in the fuel they burn.

Engine manufacturers have developed specific fuel gas specifications limiting the contaminant content within the fuels burned.

<u>Carbon Dioxide</u>: will affect engine performance and emissions in many ways. During combustion, water vapor and CO_2 can form carbonic acid that can attack certain engine parts and pollute engine oil.

However, it is generally not considered economical to remove CO_2 from engine fuel [16, p. 4-24].

<u>Hydrogen Sulfide</u>: engine manufacturers recommend that H2S be limited to under 10 ppm or 0.001% by volume. Caterpillar recommends that H_2S level must be lower than 47.5 µg/BTU to account for the gas BTU level.

<u>Volatile Organic Compounds</u>: LFG can also include volatile organic compounds (VOCs) and chlorofluorocarbons (CFCs) which, when burned in an engine, form hydrochloric acid and hydrofluoric acid which can corrode engine components.

Caterpillar recommends keeping chlorine and fluorine levels below 40 µg/BTU.

<u>Water vapor</u>: gas water vapor combines with contaminants to form organic acids and carbonic acid which contaminate engine oil. Caterpillar recommends that moisture content be kept below 115 lb/MMSCF.

<u>Particulates</u>: silica particulates should be kept below 0.4 microns to prevent abrasion.

There is evidence that engine manufacturers are engineering contaminant-resistant engines; there is also evidence from the field that unexpectedly high maintenance engine costs occur when raw landfill gas (having undergone moisture removal process) is fed to engines.

Information coming from Europe in the course of recovering biogas from collective organic waste treatment suggests that regular contracted engine maintenance keeps breakdowns in check.

It is generally recognized the all biogas conversion systems will require some form of gas cleanup.
While the removal of water and H_2S are the only cleanup required for many applications, more complex techniques may be required for more complex applications [15].

H_2S content is usually less than 1%. Corrosion may occur however, especially if the biogas is compressed. Biogas can typically be combusted in a burner or engine without causing a SO_2 problem.

H_2S levels in biogas are typically as follows:

- Animal waste: 100-1,000 ppmv
- Industrial waste water: 100-10,000 ppmv
- Municipal sludge: 1,000-20,000 ppmv
- Landfill: 100-20,000 ppmv

CO_2 reduces the heating value of biogas; removing the CO_2 will increase its heating value. High BTU gases of pipeline quality require the removal of sulfides, CO_2 and water.

Gas Production in anaerobic Digestion

The anaerobic process using methane fermentation is temperature sensitive; therefore temperature control is necessary. Most process are maintained at approximately $95^{0}F$ ($35^{o}C$) and must be heated. Methane produced can be used to heat the system if the organic concentration of the wastewater if sufficiently high [2, p.322].
Gas production varies with temperature; it ranges from between approximately 6.0 and 7.2 ft^3 of gas per lb volatile solids added, or 7.0 to 15.1 ft^3 gas per volatile solids digested [2, p. 192].
The heat value of sewage gas is 584 to 646 BTU/ ft^3.

Low-solids and high-solids anaerobic digestion

Low-solids anaerobic digestion is a biological process in which organic wastes are fermented at solids concentrations equal to or less than 4 to 8 percent.
This low-solid process is used in many parts of the world to generate methane gas from human, animal and agricultural waste as well as from the organic fraction of MSW.
The process occurs in 3 steps: receiving and preparation; addition of moisture and nutrients, pH adjustment, heating; capture, storage and separation of the gas components CH_4 and CO2; dewatering and disposal of the digested sludge.

Gas production

8 to 16 ft^3/lb (0.5 to 0.75 m^3/kg) of volatile solids destroyed.
High-solid digestion is a newer technology; it is similar to low-solid digestion except that total solid content is about 22%. This process requires less water and produces higher gas volumes per unit.
Beginning 1992, high-solid anaerobic digestion installations were operating in Europe, while several high-solid process were in development in the United States.

As of 1992, neither technology were commercialized for energy recovery [1, p. 705]

Sludge, combustion of

Multiple technologies are available for sludge combustion.
One example is the multiple hearth incineration technology manufactured by Zimpro, Wheelabrator Incineration, C-E Raymond. This technology features the combustion of sludge through successive chambers with varying temperatures from 300°F to 1,800°F. This is the most widely used technology in the United States (350 installations). Waste heat is made available for power generation.

1.2 Characteristics of sewage digester gas - biogas

- Results of a mailing survey directed at Ohio WWT plants

To investigate and record the characteristics of biogas produced by anaerobic digestion within WWT plants, AplusB sent a questionnaire to 64 of the WWT plants servicing the largest metropolitan areas in Ohio.
Replies totaled 6 in all; within these 6 replies, only one plant indicated the characteristics of the biogas produced; results are listed below:

Municipal WWT Plant biogas characteristics

Characteristics	Plant data
Methane	65.74%
Carbon Dioxide	33.99%
Nitrogen	0.27%
Oxygen	0%
Water	6%
Hydrogen Sulfide	190.83 ppm
Aromatics	
Argon	0%
Carbonyl sulfite	0.23 ppm
Sulfur dioxide	0%
Ethanol	0.16 ppm
Dinethyl sulfide	0%
Organo-halogens	
BTU gross saturated	657
BTU gross dry	699

1.3 USA Companies associated with bio digesters

- Approximately 190 of such Companies are listed[4] in Appendix A and B of reference [16]: "The Handbook of Biogas Utilization" produced for the Southeastern regional Biomass Energy Program, Tennessee Valley Authority, Alabama, by Environmental Treatment Systems, Inc., Atlanta.
- **Table A** (page 27) summarizes the leading anaerobic digestion processes in USA and Europe

1.4 Investigate how WWT sewage gas (biogas) is currently utilized in the USA

The **City of New York's** North River WWT Plant (one of 14 plants) where wastewater flows into the North River plant from an 11 mile-long intercepting sewer that extends along Manhattan's west side. Upon entering the plant, the wastewater first passes through upright bars that remove large items, including rags, sticks, newspapers, cans and other debris. The trash is automatically scraped from the bars and later transported to a landfill. Five main sewage pumps lift the wastewater to the surface level primary settling tanks. The flow of the water is slowed, allowing the heavier solids to settle on the bottom and

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[4] These lists are not appended to this report

The flow of the water is slowed, allowing the heavier solids to settle on the bottom and the lighter materials to float. Oil and grease are skimmed from the top of the tanks and the heavy solids, called "primary sludge," are scraped off the bottom for further processing.

The partially treated wastewater then flows to the secondary treatment system. Secondary treatment is called the "activated sludge process," because air and "seed" sludge from the plant treatment process are added to the wastewater to break it down further. Air pumped into five, 30-foot-deep aeration tanks stimulates the growth of oxygen-using bacteria and other organisms that consume most of the remaining organic materials that pollute the water. The aerated wastewater then flows to 16 final settling tanks, where heavy particles and other solids again settle to the bottom. Some of this sludge is recirculated back to the aeration tanks as "seed" to stimulate the treatment process. The remaining solids are removed and join the primary sludge for further processing in sludge-handling facilities. The wastewater is disinfected with sodium hypochlorite. The treated wastewater or effluent, is then released into the Hudson River.

The **City of Chicago** Water Reclamation Plants – which include some of the largest WWT plants in the world: the Stickney Plant treating 781 million gallons a day, and the Calumet plant treating 262 million gallons a day.

The reported digester gas production in 1998 for the Stickney plant totaled 1,736,000 x 10^3 SCF or 4.76 MMSCFD on average. Biogas production varies seasonally depending on a variety of factors such as: incoming solids, volatile content, etc.
The reported digester gas production in 1998 for the Calumet WWT plant totaled 324,000 x 10^3 SCF or 0.90 MMSCFD.

Stickney plant: the biogas produced in this plant satisfies 95-97% of its energy requirements for heating and digesters heating. Biogas fuels a 3 MW turbine which operates 8 months/year. Excess biogas is flared, though infrequently. In 99, only 37 MMSCF of natural gas was purchased to meet boiler needs.

The H2S content of the biogas is reduced through the "Sulfa Treat" process to reduce concentration to less than 10 ppm.

CO2 is not separated from the methane, nor is it used in any part of the plant.

Characteristics of the digester gas: 60 to 65% methane, 30 to 35% CO_2 and trace gases such as H2S, N, etc.; H2S concentration varies between 200-500 ppm.

Local sewers from each of the 125 municipalities within District boundaries connect to interceptors which gather the wastewater and convey it to one of the treatment plants. Here it undergoes a number of cleaning processes:

Primary Treatment removes contaminants by some physical mechanism:

Screens remove debris that can clog the machinery.

Wastewater flows into chambers where heavy solids such as sand and grit sink to the bottom; these solids are washed before being deposited in a sanitary landfill.

Revolving "arms" simultaneously scrape the primary (untreated) solids from the bottom and skim the grease from the top.

Secondary Treatment usually employs a biological process whereby a large population of micro-organisms help convert the remaining organic material into other forms which can be easily separated into solids and a clear liquid.

The primary effluent flows through a series of large rectangular aeration tanks which have been seeded with bacteria and other microbes . Filtered air is pumped through the liquid to enable the microbes to breathe and grow. In the constantly churning water, these microbes flourish and multiply, eating the remaining organic materials and nutrients in the wastewater.

This mixture of microbes and water flows into a secondary settling tank. The microbes, now stabilized, clump together and settle to the bottom of the tank where they become part of the organic residuals and are removed. Approximately eighty-five percent of these microbes are recycled to the start of the aeration tanks to begin the biological treatment process for the primary effluent.

The cleaned water flows out of the top of the secondary settling tank to be returned to the waterway or to the tertiary treatment process.

Tertiary Treatment is only required when the final effluent must be so clean that 95% or more of the contaminants must be removed by wastewater treatment.

Disinfecting to destroy bacteria which can cause disease. The reclaimed water has more than 95% of the impurities removed and can be deposited into a river or stream without any adverse environmental impacts. This "effluent" is often cleaner than the water of the stream. The entire process from the time wastewater reaches the treatment plant to the time it is cleaned and "reclaimed" takes less than 12 hours.

Sludge treatment

Sludge produced by primary and secondary treatment is approximately 97 percent water, and must be concentrated for further processing. It is sent to thickening tanks for a period of up to 24 hours, where it settles to the bottom. The water that remains is directed back to the aeration tanks for additional treatment. The thickened sludge, which is about 96 percent water, is then placed in anaerobic digesters and heated first to 95°F. This stimulates the growth of anaerobic bacteria. Biogas byproduct of the digestion process is used as fuel in certain plant operations.

Converting sludge into biosolids

After digestion, the sludge is dewatered; this reduces the amount of water the sludge contains, producing a moist, soil-like substance called "biosolids" that is easier to handle. Because North River has no dewatering facilities, sludge from the plant is transferred by boat for dewatering at the Wards Island wastewater treatment plant, the site of one of the City's eight dewatering facilities.

After dewatering, all of the City's biosolids, including those generated at North River, are recycled and reused. and convert into biosolids that are environmentally safe to be used as fertilizer products or directly applied onto land to enrich nutrient depleted soils. North River's biosolids are either thermally dried into fertilizer pellets, composted, or alkaline stabilized into a product that resembles soil and is used as an agricultural liming agent

The **City of Boston** Water Reclamation Plants

The centerpiece of this project is the primary treatment plant, which provides sewage treatment for metropolitan Boston and its surrounding forty-two communities.

This plant has a 1.3 billion GPD capacity and replaces the existing facilities on Deer and Nut Islands. Sewage is sent to Deer Island where it is pumped via two half-mile long, eighty five foot deep tunnels with a pipe diameter of eleven feet to remove large debris and grit.

The wastewater then travels to the primary clarifiers where the heavier suspended solids settle.

The Deer Island facility has four clarifiers which are stacked in pairs of two to conserve space. The sludge that settles in the bottom of the clarifiers, as well as the floating grease which rises to the top, are removed and sent to **digesters** to undergo a ten to twenty-two day stabilization process. Digesters are giant "eggs" (100 feet high and 90 feet in diameter) with a volume of three million gallons of wastewater.

Approximately fifty tons of sludge per day undergo the digestion process of mixing and heating. Digestion cuts the volume of sludge in half and reduces the bacteria that causes odor and disease. These state-of-the-art "eggs", which are used almost exclusively in Europe and Japan, are efficient because the superior shape eliminates "dead spots" that can cause solid build-up, and/or temporary shutdown. Also, the decreased surface area of the egg design allows for less scum buildup. A unique characteristic of the egg digesters is that they are self-cleaning and eliminates the smelly cleanup process that leaves conventional digesters with interrupted service.

Construction costs of the digester "eggs" were expected to be as much as thirty percent higher than conventional tanks; in the long run, these ovoid tanks are less expensive to operate and maintain.

The digester complex also contains two storage tanks that will hold the treated sludge, which was formally dumped into the Boston Harbor, until it is ready to be transported to the Fore River Staging Area in Quincy, Massachusetts.

Treated sludge is converted into fertilizer pellets to be sold nationwide. Approximately 1.4 million gallons of sludge will be transported to Quincy per week. Under the current system this treated sludge is transported by barge; however a five mile tunnel from Deer Island to Fore River is in the process of being built.

Primary Data

- Anaerobic digestion is in use.
    - The first phase of the Deer Island WWT facility came on line in 1995
    - The second phase came on line in September 1997
    - The project is nearing completion in 2000
- The amount of gas produced is 170,000 SCF/hour – 24 hours per day, 7 days per week. Approximately 4.08 MMSCFD.
- Gas is not analyzed for content although it is assumed to be approximately 2/3 CH_4 and 1/3 CO_2
- Tests for H_2S contents are held weekly; the target content of H_2S is below 1,000 PPM which it usually is; occurrences of higher H_2S content are noticed, perhaps due to unusually dry weather. Ferrous chloride is added which brings the H_2S content down to 100 PPM or less before the gas is burned to supply energy to a 4 MW steam generator which supplies part of the energy used at Deer Island.
- If the pH needs to be adjusted, sodium bicarbonate is used
- Biogas is piped to a 4 MW steam turbine which supplies part of the energy needs for Deer Island
- The biogas is not cleaned or separated into its constituents; water is however removed. A process to separate the CO_2 from CH_4 was considered but not adopted. It was thought that removing the CO_2 from the gas would increase efficiency and make it easier to maintain pipeline pressure to the generator.
- A process to remove particulate matter from the biogas is installed, however very little particulates in the gas are found.
- About 90% of the biogas is used in the generator. The remainder is flared. All of the biogas could be used, except that there is difficulty in maintaining the proper pipeline pressure if all the biogas is used.
- A small portion of biogas is used to power a fuel cell. For more information about the fuel cell and how the gas may have to be cleaned to be used in the fuel cell contact Dan Parry at (617)-539-3610.

Other information

- The facility was designed by Camp Dresser McKee and constructed in part by the Electric Boat Division of General Dynamics. Contact for information about Camp Dresser is Jim Small - phone: (617)-252-8403
- CO2 is not added to the process at any point.

The **East Bay Municipal Utility District** (EBMUD) – Oakland, California

Primary Data

- An anaerobic digester is in use
- The amount of biogas the digester produces was not known
- However, the characteristics of the biogas are approximately:

CH4	63% to 64%
CO2	approximately 34%
Sulfur	"very low"

Biogas usage

- The biogas is piped to a co-generation facility operated by three engines
- The biogas is not cleaned or separated; some "drip traps" in the pipe collect water
- The co-generation facility produces electricity which in turn heats water which is used to heat the digesters
- About 35-40% of the energy needs of the WWTP are derived from the biogas
- 100% of the biogas is used in the co-generation facility. If there is more gas produced than can be used in the co-generation facility, it is flared - this is not a usual occurrence

The **Point Loma Wastewater Treatment Plant** – San Diego, California

Put into service in 1963, the Point Loma Wastewater Treatment Plant treats more than 190 MGD of wastewater from a 450 square mile area.

Wastewater entering the plant first passes through screens which act as giant rakes to remove materials that would interfere with the treatment processes further along. These materials are a diverse assortment of paper and plastic products, vegetable matter, etc.

The wastewater then flows into grit removal tanks, where heavy inorganic particles such as sand, cinders, coffee grounds and eggshells settle to the bottom.
Following grit removal, sedimentation tanks remove settleable organic solids and floating materials like grease and oil. With the help of chemicals like ferric chloride and organic polymers, waste particles bond together in large enough mass to settle out. At this point, approximately 80 percent of the total suspended solids in the water have been removed. The treated wastewater, or effluent, is now ready to be discharged to the ocean through a deep ocean outfall.

The City of San Diego received a waiver from the secondary treatment requirements of the Clean Water Act in November 1995.

Through a combination of factors, including industrial source control, advanced primary treatment of wastewater, a deep ocean outfall and comprehensive monitoring, the Environmental Protection Agency and the Regional Water Quality Control Board agreed that the Point Loma Wastewater Treatment Plant fully protects the ocean. This decision saved ratepayers an estimated $3 billion. A $200 million upgrade of the 40-acre Point Loma site is currently underway.
In 1993 the City completed a two-mile extension to the original 2.5-mile-long outfall pipeline that carries treated wastewater to the ocean floor from the Point Loma plant. The original 2.5-mile-long pipeline, which followed the ocean floor to a depth of 220 feet, was extended to 4.5 miles and a depth of 320 feet, making it one of the longest and deepest ocean outfalls in the world. A Y-shaped diffuser pipe further extends from the end of the west-facing outfall in a northerly and southerly direction by about 2,500 feet to better disperse the treated wastewater.

Additional construction includes two new digesters to accommodate higher volumes of wastewater, a sludge pumping station - a critical link to the Metro Biosolids Center - an upgrade of the cogeneration facility that turns methane gas, generated by the facility's digesters, into electricity, and improvements to the odor control system.

The City of Columbus, Ohio

Primary Data

- Anaerobic digestion is used in one of two WWT treatment facilities in the Columbus WWT plant. The second treatment facility will probably be converted to anaerobic technology in the next few years. It is a one-stage mesophilic process.
- The current digester produces about 800,000 SCFD.
- The characteristics of the gas are approximately 65% CH_4 and 33% CO_2. There is no consistent or detailed analysis of the biogas generated.
- The process does not require CO_2 .

Biogas usage

- Some of the biogas is partially dried and provides heat for in-plant use, primarily for the incineration of the leftover solids.
- The balance of the biogas is flared.
- There seems to be no governmental agency concerned as to the disposition or characteristics of the biogas produced.

Miscellaneous

- Plant management believes that most future WWT plants will use anaerobic digestion to reduce quantities of solids (sludge) produced and to stabilize these solids so that they can be handled safely and efficiently. These solids can then be used for agronomic purposes.
- Characteristics of biogas resulting from anaerobic digestion should be relatively constant, regardless of the characteristics of the wastewater being treated.

- There are three major types of anaerobic processes: a low temperature mesophilic process that operates in the 95 to 98 degree Fahrenheit range, another one-stage process that operates at a slightly higher temperature range, and a two stage process; in the latter, the first stage produces 60 to 65% CO_2, While the second stage produces primarily CH_4.

The **City of Cincinnati**, Ohio

Primary Data

- An anaerobic digester is in use; it is planned to remove it in the future
- This digester produces between 600,000 and 800,000 SCFD of biogas
- Biogas characteristics are analyzed only in relation to its BTU and sulfur contents; this information was not available when we called the Plant. Other biogas characteristics provided were:

CH_4	60%
CO_2	38%
N_2	0.7%

Biogas Usage

- The design procedures called for the biogas to be dried and partially used in-plant to heat the sludge and incinerate leftover solids.
- Currently all biogas is flared.

Miscellaneous information

- Because of high maintenance costs, the digester will be closed in the future. It is also believed that outright incineration of the raw waste will provide for better odor control.
- Adjustment of the pH is performed by using ferrous chloride purchased from a neighboring manufacturing plant

The City of North Royalton, Ohio WWT Plant – Energy recovery

This newly built plant expansion (1999) was designed to accommodate the requirements of the community it serves for the next 20 years. Current total plant capacity reaches 3.3 million gallons a day.
On the solid side, dewatered sludge is processed into an open bay composting system. The nutrient rich compost will be marketed to gardeners, homeowners and businesses.

The **West Point Treatment Plant, King County**, Seattle, WA

Six digesters are in operation in that Plant.
The daily volume of biogas generated averages 2.6 MMSCFD.
Biogas is recovered for in-plant energy use.

1.5 Carbon dioxide usage in *industrial* and *municipal* WWT Plants

A significant number of ***industrial*** WWT plants in the USA purchase CO2 for their internal use.
On the other hand, we have found no evidence of ***municipal*** WWT plants purchasing CO2 for their internal use.

1.5.1 Carbon Dioxide Usage

Reportedly, carbon dioxide is added to high pH, lime-treated water to lower its pH [4, p.63]. [Note: Recarbonation is a term used to describe the process of adding carbon dioxide to lime treated water.]
There are several reasons to adjust pH: optimizing the action of coagulants (some of which are more effective in a narrow pH range); placing the water in calcium carbonate equilibrium to avoid problems of deposition of calcium scale; maintaining the adsorptive capacity of filters.
The source of CO2; in advanced WWT plants, will usually be stack gas from either a lime recalcination furnace or sludge incineration furnace; this gas should be passed through a scrubber.

Other sources would be commercial liquid carbon dioxide, or the burning of natural gas, propane, butane, kerosene, fuel oil or coke.

Commercial liquid carbon dioxide has been increasingly used for water softening plants in recent years (because of its decreasing costs).

Even in optimum conditions however, the cost of delivering liquid CO2 to WWT plants (being highly dependent on delivery distance) is considered to be high. Its advantages however are: flexibility, ease of control, high purity and efficiency, shorter piping required.

Liquid CO2 is delivered by 10 to 20 tons tank trucks or railcar shipments of 30 to 40 tons. Some suppliers will lease the tank cars to be used as storage.

1.5.2 Commercial Sources of CO2

Informal conversations with a world wide supplier of industrial gases revealed that CO2 is extensively sold to ***industrial*** WWT plants.

Examples of ***industrial*** WWT plants would be those found at bottling plants (which use caustic wash water), animal husbandry plants which use caustic cleaners for neutralization or miscellaneous industrial and other chemical plants which, as a result of their process, generate high pH streams. These streams require pH adjustments. The volume of CO2 required by these plants is relatively "small" and do not justify the investment in a specific plant to satisfy their CO2 requirements.

An interesting comment made was that the CO2 itself and every piece of equipment used in the distribution process of liquid CO2 is kept, by the industry, at "food grade" purity level.

Also interesting is the fact that the minimum capacity of smaller commercial CO2 plants would in the 100 ton per day range; the Acrion technology feasibility range matches this tonnage.

The supplier of industrial gases indicated however that new sources of CO2 might be of interest provided that it is located in a geographical area such that delivery costs from these new sources are lower than those currently incurred to deliver from a perhaps distant CO2 source.

Major sources of CO2 in the Midwest are located within economical distances from the locations where it is to be used. Such is not the case however in California and the north east coast of the USA.

There appears, therefore, to be a market niche for new sources of CO2; to be competitive, however, the purchase price that industrial producers of CO2 would be willing to pay would have to be in the $5 to $15 per ton range. Purity range in excess of 99% would be required. Additionally, it was mentioned by an industrial gas producer that removing impurities like sulfur or hydrocarbons (e.g. benzene) which might be contained in the CO2 is considered difficult and thus costly, adding perhaps $5 per ton to product costs.

2) Well informed WWT plant equipment vendors believe that, to the best of their knowledge, *municipal* WWT plants do not use CO2 in their process.

In fact, a survey among 64 *municipal* WWT plants in Ohio (which yielded 10 % responses) indicated no requirement for CO2 either current or in the foreseeable future.

1.6 Addendum: Energy Recovery from Sewage Treatment Through Methane

A recent survey of sewage treatment plants in the Southeast using anaerobic digestion with methane recovery has documented significant energy and environmental benefits.

The survey, by Wander Associates of Alexandria, Virginia, was made to identify and define the use of anaerobic digestion with methane recovery in the region. Ten plants were selected for a detailed analysis of their technology and economics. Most of the plants did not differentiate the costs associated with the anaerobic digestion system from the general plant operations, making cost data difficult to calculate. Seven had historical records showing annual savings of $67,000 to over $700,000 in avoided energy purchases. Another had insufficient records to establish a quantity but the operator believes the savings are significant. Two others lacked sufficient data to make an estimate. Wander Associates estimated that all but one of the plants could demonstrate annual savings of $5,000 to over $600,000 per Mgal/day of sewage treated.

Four plants had sufficient data to make an estimate of probable initial capital costs, which ranged from $30,000 to $160,000 per Mgal/day of sewage treated. The arithmetic mean for the four was $82,000 per Mgal/day. Payback periods of 4.3 to 15 years could be calculated for three of the plants. The four plants could also indicate that annual operating and maintenance costs ranged from about $300 to $8,700 per Mgal/day.

The effective technology and favorable economics of these plants illustrate the advantages of methane recovery and use as well as the under-utilized potential of methane recovery. Reportedly, there are about 3,000 sewage treatment plants in the Southeast with capacities of over 1 Mgal/day with the potential for methane recovery and use.

Sewage treatment plants process large volumes of liquid containing 1 to 2 percent solids by forced aeration in large basins, the "activated sludge" process. Solids are removed before aeration and the excess sludge produced during aeration is removed for disposal. When used, the anaerobic system is used to reduce the sludge volume and sometimes make use of the biogas.

Typically the biogas produced from digestion of sewage sludge contains about 50 to 60 percent methane, 35 to 45 percent carbon dioxide with traces of hydrogen sulfide and nitrogen. It is essentially a low-Btu natural gas with a heating value of 500 to 600 Btu/1000 cf. About 25 to 43 ft^3 of gas can be produced per pound of volatile solids destroyed.

Using the biogas to directly fuel the aerobic digestion blower engines or indirectly power the blower motors by generating electricity, is the most common major usage. Blower energy is the major operating expense in sewage treatment plants. Substantial savings in fuel oil and electricity were obtained in all cases. Heating the digesters using hot water boilers was also a major use.

The plants evaluated range in capacity from 120 to 9.5 Mgal/day of raw sewage. All used a two-stage completely mixed system with sludge recycle to increase the microorganism concentration, typically using digesters of about 1 million gallons. This is the simplest of anaerobic digestion systems. The digesters are maintained in the mesophilic range of about 95°F. Usually the digester feed contains about 3 to 5 percent weight total solids, of which 50 to 80 percent is volatile solids. The volatile solids reductions are typically about 50 percent and about 15 cubic feet of gas is produced per pound of volatile solids destroyed. Gas production for the plants ranged from 29,000 to 746,000 ft^3/day with an average production of 254,000 ft^3/day for the ten plants. The biogas is usually treated by passing it through iron sponge to remove hydrogen sulfide before being used to fuel blower engines and engine generators. At the Plantation, Florida plant, the gas is purified by scrubbing in a tower and is compressed for use in seven city vehicles as well as providing power for an engine generator.

In a typical example, the Muddy Creek plant in Winston Salem, North Carolina, processes 12.7 Mgal/day of sewage. Raw sludge is removed in primary clarifiers and activated sludge is removed from the aeration basins. The combined sludges contain 3.7 percent total solids, of which 82 percent are volatile solids. The sludge flow rate is about 100,000 gal/day. Sludges are pumped to four 100-foot diameter insulated concrete digesters 40 feet high, about two-thirds below ground. The digesters have gas holding covers capable of storing 14,000 ft^3 of gas at 9 inches of water pressure. Three normally act as primary digesters and one as a secondary digester although piping allows other configurations. The digesters operate at 98° F, have a hydraulic retention time of 82 days, destroy about 52 percent of the volatile solids in the feed, and produce 11.2 ft^3 of biogas per pound of volatile solids.

The biogas fires three 4.8 MBtu/hr boilers, and fuels two 800 HP, 11,500 ft^3/min engine-driven blowers. The boilers, supplemented by 12 gal/day of fuel oil, are used to provide hot water for heating the digesters and buildings.

The digesters require one to three boilers depending on the season. The blower engines are operated on biogas for 12 hours per day during the utilities' peak demand period. During off-peak periods they are driven by electric motors. Plant personnel perform most of the maintenance and repairs.

The annual savings are $46,000 in boiler fuel and $103,000 in blower electricity. According to plant historical data, annual operating and maintenance costs are $83,000, for a net annual saving of $67,000.

Anaerobic digestion is an established technology long used by a few sewage treatment plants (one of the plants surveyed, the Northside plant in Durham, North Carolina, began operation in 1933). Its use has been slow to expand, however, and until recently federal and state policies and programs have not provided effective information and support.

Several factors have given new impetus to its use. Rising sludge disposal costs have increased the desirability of reducing sludge volumes, which can be accomplished by anaerobic digestion. Also, pressure to control costs of treatment, which largely involves purchased energy, are increasing. Utilizing methane is a practical way of reducing this cost.

Anaerobic digestion is also environmentally sound if the gas is used. Methane is a powerful greenhouse gas that is addressed in the Energy Policy Act of 1992. It is estimated that the climate change caused by methane is 20 to 60 times as large as carbon dioxide. According to the Intergovernmental Panel on Climate Change, wastewater treatment produces 20 to 25 million metric tons of methane per year worldwide, 5 percent of the total emissions. By capturing the methane, even though more is produced by the process, methane emissions from the sewage plant and from sludge disposed of in landfills is reduced. The methane also replaces fossil fuels, thus mitigating emissions of carbon dioxide from fossil fuels.

* * *

Chapter 2: Biodigestion technologies advances in Western Europe

Advances in anaerobic digestion systems and treatment of biogas for productive use in Western Europe was investigated.

2.1 Introduction

In Western Europe, the result of the anaerobic fermentation of organic matter is mostly called "biogas".

The fermentation process can be either spontaneous, as in landfills, or artificially generated in **"biodigesters"** where liquid and solid matter is submitted to the action of bacteria.
The resulting "Biogas" is a mixture of methane, carbon dioxide, nitrogen, water, hydrogen sulfide, heavy metals and other contaminants.

The most common European biodigester technology recirculates "biogas" to produce a desired mixing action.

Several household organic waste methanization technologies are currently offered in Europe [6]:

- Kompogas, Switzerland
- Dranco, Belgium
- BTA, Germany
- Avecon, France
- Biocel, France
- Valorga, France is an integrated process, well known in Europe. Its biodigester phase does not include mobile parts. A Valorga installation, located in Amiens, processes 86,000 tons of organic waste yearly. Initially, the biogas: produced was first injected into the pipeline grid (after contaminant removal); currently, this biogas is delivered as boiler fuel to local industries. The first installation did not meet expectations. The compost produced included debris that had not been removed by the upstream screening phase. Later installations in the Netherlands, Germany and Switzerland have corrected this situation. High quality compost is sold; methane is injected into pipeline gas or used as an energy source to produce electricity.

2.2 Characteristics of "Biogas" in Western Europe – Energy recovery usage

France [6]

Biogas Characteristics	Valorga technology	Effluents from distilleries	WWT Plant Valenton
Methane	60%	68%	65%
Carbon Dioxide	33%	26%	33%
Nitrogen	1%	1%	
Oxygen	0%	0	
Water	6%	5%	
Hydro Sulfide	100-900 mg/m3	400 mg/m3	2%
Aromatics	0-200 mg/m3	0	
Organo-halogens	100-800 mg/m3	0	

ADEME (Agence de l'Environnement et de la Maîtrise de l'Energie – *translation: Agency for Energy and Environment Management*) and GDF (Gaz de France - - *translation: national gas utility company*) are partnering to study the commercial applications of "biogas"; they are sponsoring the following projects:

a) Research Projects to select the most appropriate contaminant removal technologies,
b) Feasibility studies based on three pilot plants (landfills),
c) A "biogas to energy" guide intended for the French Municipalities,
d) Data gathering from domestic (France) and international sources to document technical, social, commercial aspects of "biogas" applications.

France operates 150 digesters installations (i.e. biodigesters) within waste water treatment plants.
The most frequently used technology in France is called "infinitely mixed"; essentially, mixing by hydraulic means or by recirculation causes the "biogas" to be produced.
Several improvements were recently introduced; such as the BTA process which separates the floating waste from heavier waste and shreds the vegetal cells, using the plant's recycled water.
More modern technologies (proposed by e.g. Propersol) increases the surface area where the bacteria can find support, thus accelerating the fermentation process considerably (a few hours instead of a few days) and generating biogas enriched in methane (up to 80%)

France's municipalities have encouraged neighboring industries to bio digest their waste jointly (e.g.: the REVICO Company, in France, biodigest waste produced by several Cognac producers).

WWT Plant, methane utilization, France

Example of methane produced in a WWT - France (Valenton, Seine)
The fermentation process at this WWT produces 38,500 cubic meters per day – or 1.3 MMSFD of digester gas.
One third of the biogas is used for digester heating purposes (which requires 35°C) and for in-situ sludge incineration purposes.
One third of the biogas fuels dual IC engines driving generators producing 30,000 kWh (32% of total plant electricity demand).
Note: engines incur high maintenance costs (US$ 210,000 per year).
The remaining gas is flared [6].

Waste to energy Projects, France, urban waste versus industrial waste

In France, waste to energy projects focused on "biogas" generated by municipal waste are loosing ground. If related Projects need to be economically viable, subsidies are required; these seem to be in short supply.
On the contrary, waste-to-energy projects focused on the methanization (anaerobic digestion) of industrial wastes, are increasing rapidly with an average of 5 new Projects being added yearly.

Projects which generate electricity are believed to be economically viable if 500 kW are produced using diesel-gas engines and 1 MW are produced using gas turbines.

Pipeline gas, from "biogas", contaminants removal, France

Gaz de France (French national gas utility) is reluctant to allow biogas to be injected into the grid. This concern is related to potential copper tubing corrosion which is linked to the presence of oxygen, hydrogen sulfide and chlorinated components.
Gaz de France views contaminants removal technologies as intricate and costly.
France believes that, given the same amount of kW produced, investments needed to build contaminant removal plants are equivalent to investments need to build utility plants.

"Biogas", from WWT, used as boiler fuel

In Bayeux, France, a WWT biodigester produces 470 m^3 or 17,000 ft^3 per day of "biogas".
Characteristics: 67% CH_4, 31% CO_2, H_2S.(eliminated by circulating over ferrous oxide).
The gas is used primarily for internal requirements of the WWT Plant and boiler fuel for HVAC and municipal swimming pool heating purposes.

Denmark

In 1984, Denmark installed its first centralized biodigester facility that collects animal waste, WWT sludge, fish industry wastes, pharmaceutical wastes. The "biogas" produced, after removal of contaminants, is sold to a local town where it is used to produce heat and electricity.

Initial investment is the equivalent of 8 million US $. Governmental subsidies equal 3 million US $ annually. The return on investment does not however account for commercial fertilizers avoidance costs. The first 10 units built were subjected to an economic evaluation to determine if subsidies could be reduced. This is not yet the case.
Total biogas volumes however has increased beyond anticipated volumes.

Originally, these plants were designed primarily for energy recovery purposes; it later became apparent that centralized biogas plants make a significant contribution to solving a number of environmental problems in the agriculture, waste recycling and greenhouse gas reduction.

At this time, Denmark is operating 20 centralized plants. Currently 75% of biomass plant input is animal manure (35 to 40 million tons annually), the remaining 25% is waste that originates from food processing industries.

Denmark is currently experimenting with mesophilic fermentation, characterized by a temperature of 55°C. This higher temperature accelerates the fermentation process and thus production of "biogas" is increased.

Denmark has initiated aggressive energy and environmental management policies which have facilitated the feasibility of these centralized biogas plants, such as:

- 50% of organic waste should be recycled by 2005,
- 1995 biogas production level to be quadrupled by 2005,
- 20% reduction of the 1988 CO_2 emission level by 2005.
- investments grants amounting to 20 to 40 % of the investment costs,
- exempting biogas, and heat from biogas, from energy tax,
- offering an investment grant of DKK 0.27 or US $0.038 per kWh of electricity produced.

Daily biogas production in these centralized biomass plants range from 14,526 to 773 cubic meters per day or 510 000 SCFD to 27 000 SCFD.
Investments range from 55 million DKK or US $8 million, to 5.8 million DKK or US $0.8 million.

Note: "Farmhouse" methanization Projects in Europe, Africa and South America have not been successful. In China, India, Nepal however, millions of "farmhouse" digesters are in operation. Initial installation in China, however, were of poor quality and did not last more than 1 or 2 years. Improvements have been made. Nepal has been more successful, thanks to the assistance from the Dutch Government.

Heavy Metals, inhibition of anaerobic digestion by

In Europe, many industrial discharges of heavy metals are ultimately dispersed to the environment via wastewater treatment (WWT) plants [5, p. 104]
The following topics are discussed:

- sources of heavy metals
- removal of metals in sewage treatment plants
- mechanism of removal of heavy metals
- toxicity of heavy metals in the anaerobic stabilization of the sludge

Methane, practically free from hydrogen sulfide, from the anaerobic digestion of solid and liquid wastes from a distillery, Europe

Laboratory research and pilot plant operations at the Technical University of Athens, Greece suggest that commercial developments of anaerobic digestion, in the mesophilic range, capable of producing large amounts of energy is only possible if the rate of anaerobic digestion is sufficiently high.
Reportedly, a biogas consisting of 93 to 97% of methane, practically free from hydrogen sulfide is produced. It satisfies the thermal energy needs of the distillery. There has been some controversy about these results.
Reported yield of biogas is 0.4 to 0.7 m^3/kg of organic waste [5, p. 66].

Contaminants, removal of

Application	Heat	Electricity	Fuel	Pipeline gas
Constituents to be removed	Water	Water	Water	water
	H_2S	H_2S	H_2S	H_2S
		Organo-halogens	Organo-halogens	Organo-halogens
			CO_2	CO_2
			Metals	Metals
				Oxygen

2.3 Biogas, produced by WWT Plants in Europe

PAQUES Anaerobic Digestion Process, The Netherlands, description of

The PAQUES Solid Waste Systems B.V. organization, in the Netherlands, has developed an anaerobic digestion process for different kinds of solid wastes. The process is a wet mesophilic digestion process (Dry matter content in the reactor of 5 to 12% and a process temperature of 30-40°C.

The process can be carried out in one reactor for relatively constant and slowly degrading waste streams (such as mixed household waste and biowaste, or in two reactors for variable loads characterized by rapidly degrading waste (such as leftovers from daily open market operations).

A full scale dual reactor plant has been in operation since 1987 in Breda, The Netherlands, processing 7,000 to 15,000 tons of unsold fruits and vegetables. Biogas produced is converted to heat and electricity in a 2 x 85 kW heat power generator.

In 1992, a pilot plant, processing mixtures of municipal solid waste, industrial organic waste and biowaste went into operation.

Data gathered from operating the Pilot Plant has been used to develop a feasibility study which addresses the case of 10,000 to 30,000 tons per year of biowaste, using the single reactor technology.

Conclusions

Investment and operating expenses are relatively low compared to other digestion processes. The investment is 900 to 1050 Deutsche Marks per ton (US$ 500 to 583 per ton) of biowaste input per year. Operational costs range from 175 to 210 Deutsche Marks (US$ 97 to 117) per ton of biowaste, per year. These figures are relevant for "turnkey" plants that include power generation and which conform to the most stringent German technical standards.

Adjustment to current cost factors need to me made to evaluate investment and operational expenses.

Biogas production ranges from 80 m^3 to 120 m^3 per ton of organic matter. The biogas includes 55 to 64% of methane.

Europe, general information on anaerobic digestion process

Europe's food and agronomic industries generate large amounts of waste slurries; the necessity of processing these slurries is increasing.
The demand for food slurries as cattle feed is decreasing; hauling and dumping waste costs are rising rapidly; landfilling of organic wastes may become forbidden in Europe in the near future.
In countries such as Switzerland and Germany, electricity produced from alternative sources are sold at prices ranging from Dfl 0.15 to 0.20 per kWh (US$ 0.075 to 0.10 per kWh).

The **DRANCO** process, Belgium, anaerobic digestion process

Organic Waste Systems, N.V., Belgium, has developed the DRANCO technology, a waste treatment process producing biogas.
Investment economics come from savings in waste tipping fees. Energy produced is about 200 kWh per ton of waste; income from energy produced is about US$ 15.00 per ton of waste.
Typical tipping fees in Europe (1999 figures) range between US$ 60.00 and 120.00.
Cost per ton of the Dranco process is around US$ 80.00 per ton.
Organic W-aste Systems has installed demonstration plants in Belgium, Indonesia, Austria, Japan and USA.
Commercial plants are in operations in Belgium (35,000 tons and 20,000 tons /year); Austria (20,00 tons /year); Germany (13,500 tons/year); Switzerland (11,000 tons/year),

* * *

Table A: leading anaerobic digestion processes in USA and Europe offering potential for energy recovery

Anaerobic digestion Process	Country	Status	Description
Sequential batch anaerobic composting (SEBAC)	USA	Experimental stage	Batch anaerobic three-stage process. Volatile acids and other fermentation products are converted to methane.
High-solid anaerobic digestion aerobic composting process	USA	Under Development	Two stage process. First stage involves the digestion of the solid content to convert the organic fraction of MSW to methane. Second stage produces humus.
Semi-solid anaerobic Digestion/aerobic composting process	Italy	Under Development	Two stage process. Anaerobic Digestion/aerobic composting. First stage involves the digestion of the solid content to convert the organic fraction of MSW to energy. Second stage produces humus.
DRANCO process	Belgium	Developed	Conversion of the organic portion of MSW to produce energy and humus-like product.
BTA process	Germany	Developed	Treats the organic fraction of MSW, including the mechanization of dissolved biogenous materials. Compost-like material is produced.
VALORGA process	France	Developed	Includes sorting unit, methane producing unit and refining unit.
VALORGA process Dranco process Kompogas Funnel Industries	France Belgium Switzerland USA		Dry continuous digestion involves a continuously-fed digestion vessel with a digestate dry matter content of 20 to 40%. The requirement for minimal water additions makes the overall heat balance favorable for operation at thermophilic digestion temperatures (54 to 55°C).
BIOCELL process	Netherlands		Dry batch digestion involving loading a vessel with Municipal Solid waste and digestate from another reactor. The vessel is sealed and left to digest naturally. Leachates are recirculated to maintain uniform moisture content and redistribute soluble substrates and methane-producing bacteria. Process is simple, but the batch treatment and post-treatment of the digestate could be inconvenient. The system was developed to treat fruit, yard and vegetable wastes.

Table B: methane as a conversion product

Process	Conversion Product	Preprocessing
Anaerobic digestion (in landfill)	Methane and carbon dioxide	None required other than placement in containment cells.
Anaerobic Digestion (low-solids, 4 to 8 percent solids)	Methane and carbon dioxide, digested solids	Separation of organic fraction, particle size reduction.
Anaerobic Digestion (high-solids, 22 to 35 percent solids)	Methane and carbon dioxide, digested solids	Separation of organic fraction, particle size reduction.

ADDENDUM: factors that influence the quantity of digester gas produced and conditions under which these factors may be adjusted to increase flow of digested gas [16, p. 2-6].

Temperature: most prevalent temperature range is within 68 –113 °F, the optimum occurs around 95°F. Thermophilic temperature (113°F -140°F) operation is also possible. However, a small fluctuation from established effective range can upset process.

Hydraulic retention time: depends on influent concentration, type of influent and temperature.

Organic loading rate: depends on system design, expressed in terms of mass of COD or VSD per volume of reactor.

Air: is to be strictly excluded as it is toxic to the anaerobic digestion process

Bacteria: dependent on waste and temperature.

Carbon/Nitrogen ration: less than 43:1

Carbon/Phosphorous: less than 187:1

pH: optimum range is near 7.0; successful range 6.0-8.0

Volatile acid: bicarbonate alkalinity should exceed volatile acids alkalinity.

Solid contents: optimum sludge solid contents is 7-9% by weight.

Toxic substances: cations and heavy metals in sufficient concentration can be harmful. In general, high concentration of halogenated organics can be harmful.

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Definitions

Anaerobic treatment	Microbiological degradation of organic matter in the absence of molecular oxygen
Biogas	Gas generated from the biological, anaerobic degradation of organic matter. Biogas is typically made up of methane and carbon dioxide that make up 90% of the volume. H2S is generated from the conversion of the sulfur compounds within the treated substrate. Trace amounts of nitrogen, hydrogen, methyl mercaptans and oxygen are typically found in biogas. Landfill gas (LFG) can have a significant number of other gases including CFC's and HFC's.
Conversion	Use of biogas in different useable energy recovery applications (e.g. direct combustion systems, engine systems, natural gas pipeline sales)
Sludge	Sludge produced by primary and secondary treatment is about 97% water and must be concentrated for further processing. It is sent to thickening tanks for a period of up to 24 hours, where it settles to the bottom of the tanks. The remaining water is directed back to the aeration tanks for additional treatment. The thickened sludge which is 96 percent water is then placed in oxygen-free digesters and heated to 96 degrees Fahrenheit. This stimulates the growth of anaerobic bacteria which react with the organic content of the sludge. Methane gas is produced together with carbon dioxide and other constituents.

Engineering Units - Conversion Factors – Equivalents

1 cubic meter	35.314 cubic feet
1 kWh	3.6 x 10^6 Joules
1 kW	56.869 BTU/minute
1 kWh	1.341 HP-hour
1 HP	746 Watts
1 metric ton LNG	624 US gallons
4 000 000 SCFD	4 720 cubic metes per hour
1 lb of organic waste totally decomposed	*12 cubic feet of biogas*
4 x 10^6 cubic feet of landfill gas	2.25 x 10^6 SCFD of methane per day, plus 88 ton per day liquid CO_2
Heat value of sewage gas	584 to 646 BTU/cubic feet
1 000 tons of municipal waste (assuming it is totally decomposed) produces, per year	72, 400 SCFD of biogas
Waste water from urban households generally include 3% of <u>total</u> solids, of which 50 to 80% is volatile solids. On average, 50% of these solids are destroyed by anaerobic digestion; 15 cubic feet of biogas per lb of volatile solids is generated by AD.	Thus 1 000 000 GD (581 000 lbs) of waste water can potentially produce: 581 000 x 0.03 x 0.50 x 0.50x15 =0.065 MMSCFD (assumes solids density = 1) These theoretical figures that need to be applied with caution.

Index

* * *

APPENDIX D

Digester Gas to LNG and Liquid CO_2

Process Flowsheets, Material and Energy Balance and Equipment Size

Hysim Process Simulator

FIGURE D-1
CONTAMINANT AND BULK CO2 REMOVAL

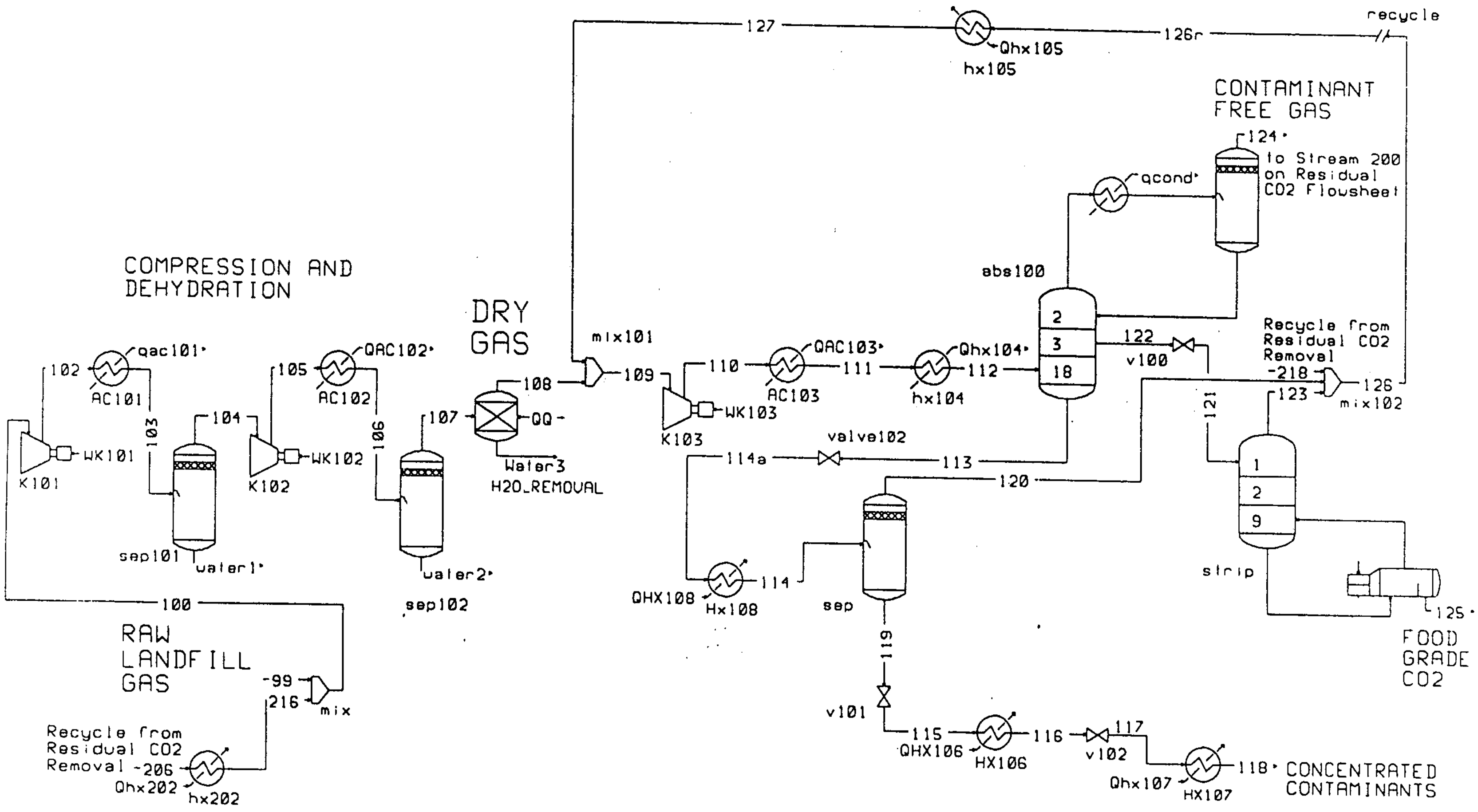

Table D-1, Material Balance for Figure D-1

Stream		99	100	102	103
Vapour frac.		1.0000	1.0000	1.0000	0.9944
Temperature	F	70.0000*	71.0984	318.6495	100.0000*
Pressure	psia	14.7000*	14.7000	60.0000*	55.0000
Molar Flow	lbmole/hr	399.4000*	430.5100	430.5100	430.5100
Enthalpy	Btu/hr	1.67842E+06	1.81220E+06	2.80569E+06	1.86530E+06

Stream		104	105	106	107
Vapour frac.		1.0000	1.0000	0.9869	1.0000
Temperature	F	100.0000	352.5121	100.0000*	100.0000
Pressure	psia	55.0000	220.0000	215.0000	215.0000*
Molar Flow	lbmole/hr	428.1106	428.1106	428.1106	422.4915
Enthalpy	Btu/hr	1.89929E+06	2.91364E+06	1.74709E+06	1.82661E+06

Stream		108	109	110	111
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	100.0000*	96.6169	318.9138	100.0000*
Pressure	psia	210.0000*	210.0000	715.0000	715.0000
Molar Flow	lbmole/hr	420.2661	541.7983	541.7983	541.7983
Enthalpy	Btu/hr	1.81921E+06	2.32289E+06	3.39757E+06	2.12604E+06

Stream		112	113	114	114a
Vapour frac.		1.0000	0.0000	0.6763*	0.2883
Temperature	F	7.0000*	-7.1979	-34.0278	-50.5966
Pressure	psia	715.0000*	715.0000	210.0000*	210.0000
Molar Flow	lbmole/hr	541.7983	73.5109	73.5109	73.5109
Enthalpy	Btu/hr	1.52799E+06	-120425.3312	69630.8359	-120425.3312

Stream		115	116	117	118
Vapour frac.		0.1035	0.9955	0.9963	1.0000
Temperature	F	-60.0000*	-40.0000*	-47.0000*	85.0000*
Pressure	psia	103.6486	98.6486	68.8866	63.8866
Molar Flow	lbmole/hr	23.7955	23.7955	23.7955	23.7955
Enthalpy	Btu/hr	-75465.6719	70370.9224	70370.9224	100918.7441

Stream		119	120	121	122
Vapour frac.		0.0000	1.0000	0.2565	0.0000
Temperature	F	-34.0278	-34.0278	-46.1838	-10.9756
Pressure	psia	210.0000	210.0000	250.0000*	701.7647
Molar Flow	lbmole/hr	23.7955	49.7154	161.9999	161.9999
Enthalpy	Btu/hr	-75465.6719	145092.2434	-277689.2632	-277689.2632

Stream		123	124	125	126
Vapour frac.		1.0000	1.0000	0.0000	1.0000
Temperature	F	-41.4384	-75.0001	-11.7634	-42.9555
Pressure	psia	250.0000	700.0000	250.0000	210.0000
Molar Flow	lbmole/hr	58.2784	306.2875	103.7215	121.5338
Enthalpy	Btu/hr	167377.1771	625765.8280	-278690.7251	353303.9577

Table D-1, Material Balance for Figure D-1, Continued

Stream		99	118	124	125
Description		LFG	Contam Gas	Clean Gas	CO2 Prod
Methane	lbmole/hr	239.2803*	0.3263	254.8686	0.0005
Nitrogen	lbmole/hr	4.3135*	0.0012	4.3540	0.0000
CO2	lbmole/hr	145.4752*	23.3813	47.0646	103.7190
H2S	lbmole/hr	0.0002*	0.0000	0.0000	0.0001
M-Mercaptan	lbmole/hr	0.0039*	0.0039	0.0000	0.0000
Refrig-40	lbmole/hr	0.0002*	0.0002	0.0000	0.0000
VinylCl	lbmole/hr	0.0015*	0.0015	0.0000	0.0000
ClC2	lbmole/hr	0.0008*	0.0008	0.0000	0.0000
Cl2-Cl	lbmole/hr	0.0098*	0.0098	0.0000	0.0000
Cl3-C2=	lbmole/hr	0.0024*	0.0024	0.0000	0.0000
Refrig-12	lbmole/hr	0.0032*	0.0028	0.0000	0.0003
Acetone	lbmole/hr	0.0137*	0.0137	0.0000	0.0000
Benzene	lbmole/hr	0.0008*	0.0008	0.0000	0.0000
Toluene	lbmole/hr	0.0273*	0.0273	0.0000	0.0000
p-Xylene	lbmole/hr	0.0098*	0.0098	0.0000	0.0000
Propane	lbmole/hr	0.0020*	0.0003	0.0003	0.0013
n-Butane	lbmole/hr	0.0020*	0.0017	0.0000	0.0003
n-Pentane	lbmole/hr	0.0058*	0.0058	0.0000	0.0000
n-Hexane	lbmole/hr	0.0058*	0.0058	0.0000	0.0000
H2O	lbmole/hr	10.2420*	0.0000	0.0000	0.0000
Total:	lbmole/hr	399.4000*	23.7955	306.2875	103.7215

Stream		109	113	122	126
Methane	lbmole/hr	288.3680	10.3288	23.1706	44.2993
Nitrogen	lbmole/hr	4.6399	0.0900	0.1959	0.3244
CO2	lbmole/hr	248.6962	63.0008	138.6309	76.9049
H2S	lbmole/hr	0.0002	0.0001	0.0001	0.0001
M-Mercaptan	lbmole/hr	0.0044	0.0044	0.0000	0.0005
Refrig-40	lbmole/hr	0.0002	0.0002	0.0000	0.0000
VinylCl	lbmole/hr	0.0018	0.0018	0.0000	0.0003
ClC2	lbmole/hr	0.0009	0.0009	0.0000	0.0000
Cl2-Cl	lbmole/hr	0.0100	0.0100	0.0000	0.0002
Cl3-C2=	lbmole/hr	0.0025	0.0025	0.0000	0.0000
Refrig-12	lbmole/hr	0.0043	0.0040	0.0003	0.0011
Acetone	lbmole/hr	0.0137	0.0137	0.0000	0.0001
Benzene	lbmole/hr	0.0008	0.0008	0.0000	0.0000
Toluene	lbmole/hr	0.0273	0.0273	0.0000	0.0001
p-Xylene	lbmole/hr	0.0098	0.0098	0.0000	0.0000
Propane	lbmole/hr	0.0030	0.0009	0.0017	0.0010
n-Butane	lbmole/hr	0.0027	0.0025	0.0003	0.0008
n-Pentane	lbmole/hr	0.0066	0.0066	0.0000	0.0008
n-Hexane	lbmole/hr	0.0060	0.0060	0.0000	0.0002
H2O	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	541.7983	73.5109	161.9999	121.5338

Table D-1, Material Balance for Figure D-1, Continued

Stream		120	123	water1	water2
Methane	lbmole/hr	10.0024	23.1701	0.0000	0.0000
Nitrogen	lbmole/hr	0.0887	0.1959	0.0000	0.0000
CO2	lbmole/hr	39.6196	34.9119	0.0002	0.0016
H2S	lbmole/hr	0.0000	0.0000	0.0000	0.0000
M-Mercaptan	lbmole/hr	0.0005	0.0000	0.0000	0.0000
Refrig-40	lbmole/hr	0.0000	0.0000	0.0000	0.0000
VinylCl	lbmole/hr	0.0003	0.0000	0.0000	0.0000
ClC2	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Cl2-C1	lbmole/hr	0.0002	0.0000	0.0000	0.0000
Cl3-C2=	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Refrig-12	lbmole/hr	0.0011	0.0000	0.0000	0.0000
Acetone	lbmole/hr	0.0001	0.0000	0.0000	0.0000
Benzene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Toluene	lbmole/hr	0.0001	0.0000	0.0000	0.0000
p-Xylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Propane	lbmole/hr	0.0006	0.0004	0.0000	0.0000
n-Butane	lbmole/hr	0.0008	0.0000	0.0000	0.0000
n-Pentane	lbmole/hr	0.0008	0.0000	0.0000	0.0000
n-Hexane	lbmole/hr	0.0002	0.0000	0.0000	0.0000
H2O	lbmole/hr	0.0000	0.0000	2.3993	5.6174
Total:	lbmole/hr	49.7154	58.2784	2.3995	5.6190

Table D-1, Material Balance for Figure D-1, Continued

Stream		126r	127	206	209
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	-42.9558*	85.0000*	-64.9602*	-81.4964*
Pressure	psia	210.0000*	210.0000	20.0000*	18.0000*
Molar Flow	lbmole/hr	121.5322*	121.5322	31.1100*	63.3577*
Enthalpy	Btu/hr	353300.4351	503682.0083	92261.8223	184337.6119

Stream		212	213	216	218
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	-44.0000*	85.0000*	85.0000*	-50.7714*
Pressure	psia	16.0000	15.0000	15.0000	210.0000*
Molar Flow	lbmole/hr	63.3577	63.3577	31.1100	13.5400*
Enthalpy	Btu/hr	203911.1985	273054.5232	133775.7108	40834.5351

Stream		water1	water2	Water3	128
Vapour frac.		0.0000	0.0000	---	1.0000
Temperature	F	100.0000	100.0000	---	-23.5032*
Pressure	psia	55.0000	215.0000	---	700.8824*
Molar Flow	lbmole/hr	2.3995	5.6190	2.2254	563.8818*
Enthalpy	Btu/hr	-33989.1166	-79525.0451	---	1.41405E+06

Stream		WK101	WK102	WK103
Enthalpy	hp	390.4545	398.6563	422.3625

Stream		qcond	qreb	qac101	QAC102
Enthalpy	Btu/hr	1.30034E+06	166377.8089	940388.1481	1.16655E+06

Stream		QAC103	Qhx104	Qhx105	QHX106
Enthalpy	Btu/hr	1.27152E+06	598046.5174	150381.5772	145836.5878

Stream		Qhx107	QHX108
Enthalpy	Btu/hr	30547.8221	190056.1707

Stream		Qhx200	Qhx201	Qhx202
Enthalpy	Btu/hr	19573.5871	69143.3243	41513.8878

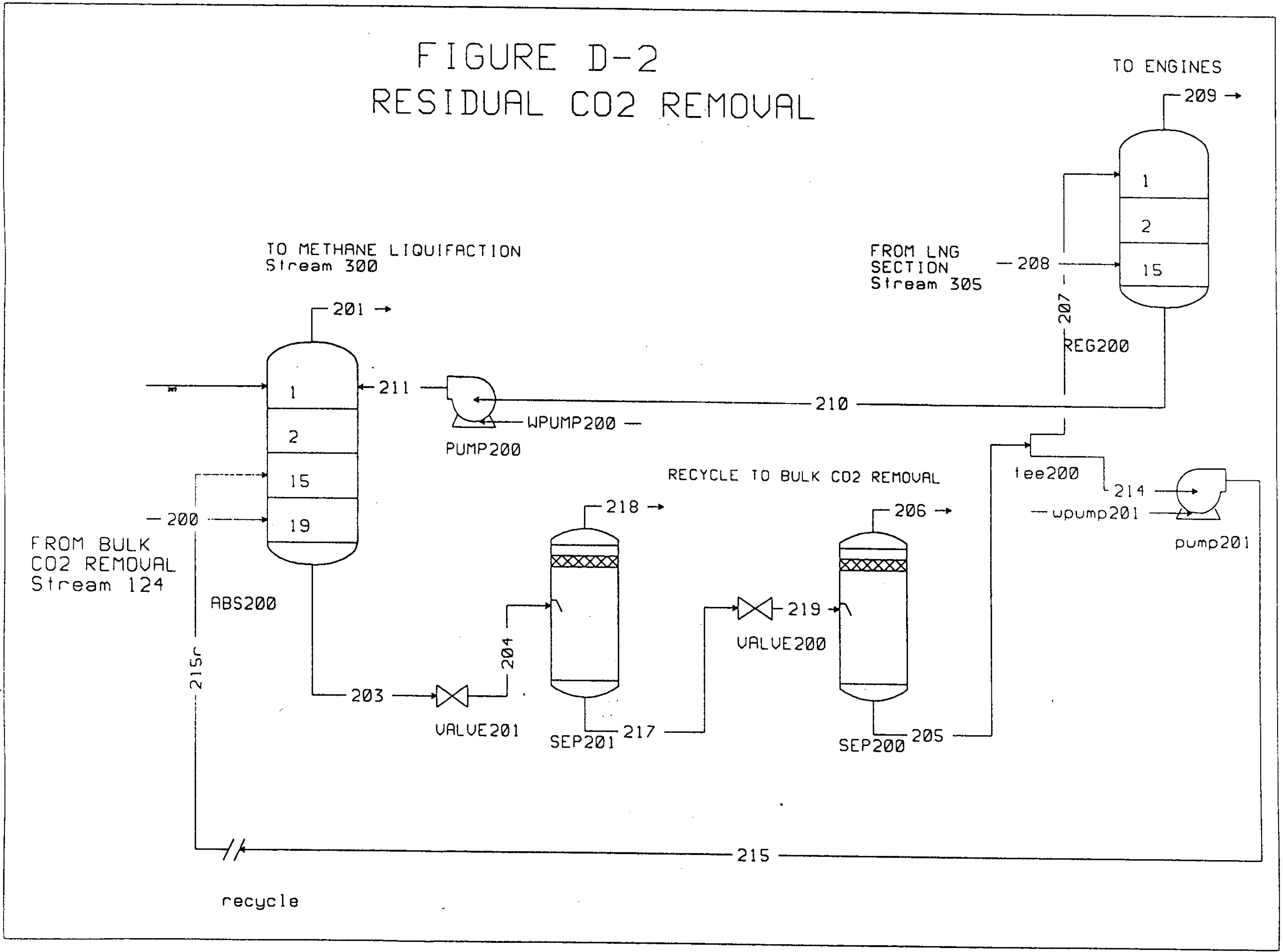

FIGURE D-2
RESIDUAL CO2 REMOVAL

Table D-2, Material Balance for Figure D-2

Stream		200	201	202	203
Vapour frac.		1.0000	1.0000	0.0000	0.0000
Temperature	F	-74.0422	-90.3007	-90.3000	-52.0494
Pressure	psia	700.0000	680.0000	690.0000	700.0000
Enthalpy	Btu/hr	640042.5983	514012.9811	-2.86579E+06	-6.54837E+06
Methane	lbmole/hr	254.3080	238.5083	0.1522	15.9972
Nitrogen	lbmole/hr	4.6766	4.6359	0.0011	0.0418
Methanol	lbmole/hr	0.0000	0.0023	157.4156	370.6404
CO2	lbmole/hr	47.1354	0.0064	0.0110	72.0977
H2S	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	306.1200	243.1529	157.5800	458.7771

Stream		204	205	206	207
Vapour frac.		0.0295	0.0000	1.0000	0.0000
Temperature	F	-50.7806	-64.9754	-64.9754	-64.9754
Pressure	psia	210.0000	20.0000	20.0000	20.0000
Enthalpy	Btu/hr	-6.54837E+06	-6.68169E+06	92369.2925	-2.83771E+06
Methane	lbmole/hr	15.9972	0.0787	4.7907	0.0334
Nitrogen	lbmole/hr	0.0418	0.0000	0.0020	0.0000
Methanol	lbmole/hr	370.6404	370.6353	0.0045	157.4090
CO2	lbmole/hr	72.0977	43.3833	26.3404	18.4249
H2S	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	458.7771	414.0973	31.1375	175.8673

Stream		208	209	210	211
Vapour frac.		1.0000	1.0000	0.0000	0.0000
Temperature	F	-80.0000	-81.4986	-92.2269	-90.3274
Pressure	psia	20.0000	18.0000	20.0000	690.0000
Enthalpy	Btu/hr	133541.0633	184390.5681	-2.88858E+06	-2.86573E+06
Methane	lbmole/hr	40.9321	40.8136	0.1519	0.1519
Nitrogen	lbmole/hr	4.1275	4.1264	0.0011	0.0011
Methanol	lbmole/hr	0.0000	0.0034	157.4056	157.4056
CO2	lbmole/hr	0.0005	18.4143	0.0110	0.0110
H2S	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	45.0600	63.3577	157.5696	157.5696

Table D-2, Material Balance for Figure D-2, Continued

Stream		214	215	215r	217
Vapour frac.		0.0000	0.0000	0.0000	0.0000
Temperature	F	-64.9754	-62.6887	-62.6885	-50.7806
Pressure	psia	20.0000	700.0000	700.0000	210.0000
Enthalpy	Btu/hr	-3.84397E+06	-3.80862E+06	-3.80863E+06	-6.58933E+06
Methane	lbmole/hr	0.0453	0.0453	0.0453	4.8694
Nitrogen	lbmole/hr	0.0000	0.0000	0.0000	0.0020
Methanol	lbmole/hr	213.2263	213.2263	213.2271	370.6397
CO2	lbmole/hr	24.9584	24.9584	24.9577	69.7236
H2S	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	238.2300	238.2300	238.2300	445.2348

Stream		218	219
Vapour frac.		1.0000	0.0699
Temperature	F	-50.7806	-64.9754
Pressure	psia	210.0000	20.0000
Enthalpy	Btu/hr	40967.0008	-6.58933E+06
Methane	lbmole/hr	11.1278	4.8694
Nitrogen	lbmole/hr	0.0397	0.0020
Methanol	lbmole/hr	0.0007	370.6397
CO2	lbmole/hr	2.3741	69.7236
H2S	lbmole/hr	0.0000	0.0000
Total:	lbmole/hr	13.5423	445.2348

Stream		WPUMP200	wpump201
Enthalpy	hp	8.9821	13.8961

FIGURE D-3
METHANE LIQUEFACTION

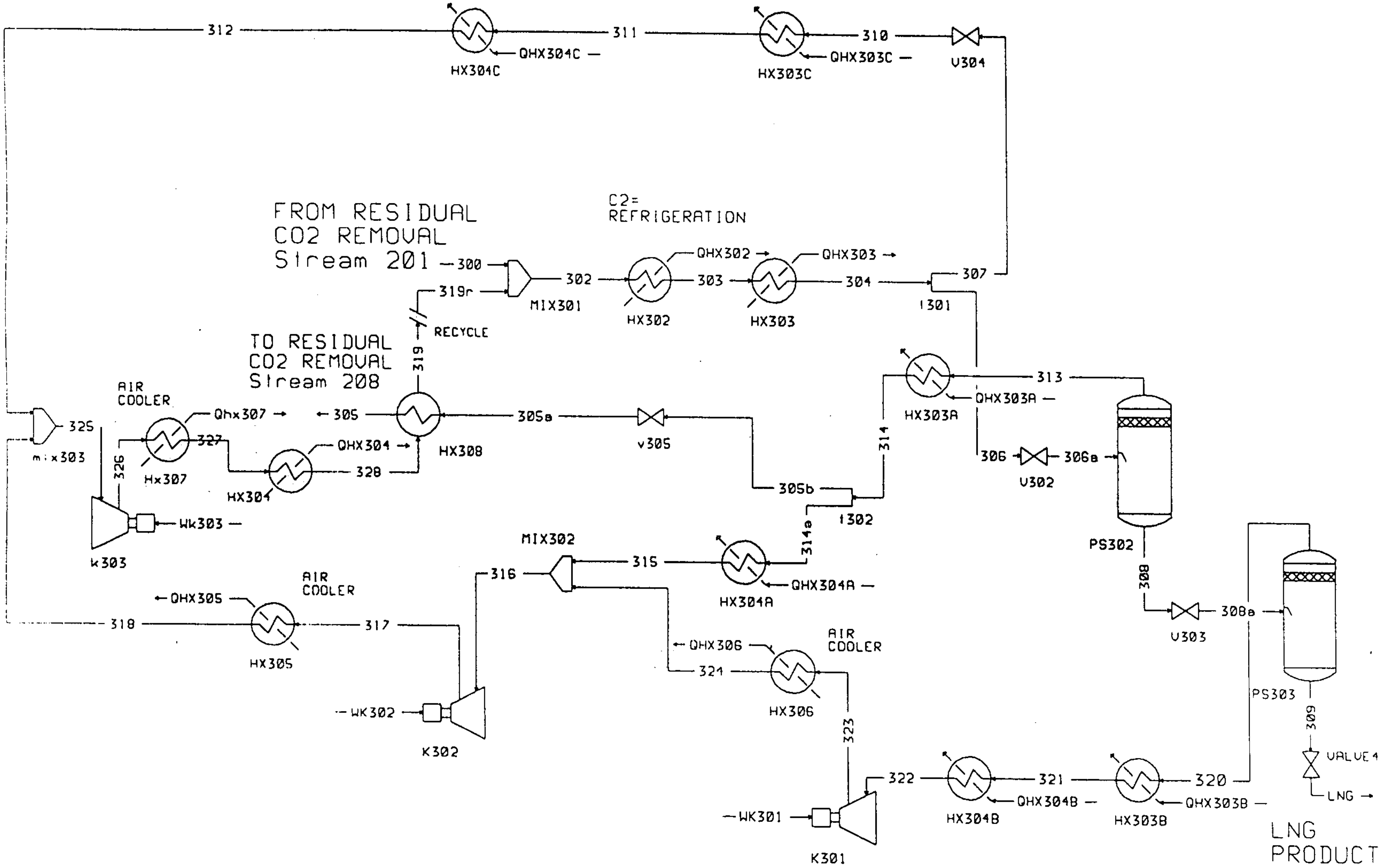

Table D-3, Material Balance for Figure D-3

Stream		300	302	303	304
Vapour frac.		1.0000	1.0000	0.0000	0.0000
Temperature	F	-90.3300*	-85.0706	-124.7006*	-172.8749*
Pressure	psia	680.0000*	680.0000	670.0000	659.9999
Enthalpy	Btu/hr	508032.8932	963868.7588	212447.6823	-290034.4908
Methane	lbmole/hr	238.5189*	429.8281	429.8281	429.8281
Nitrogen	lbmole/hr	4.6196*	12.9335	12.9335	12.9335
Methanol	lbmole/hr	0.0022*	0.0032	0.0032	0.0032
CO2	lbmole/hr	0.0122*	0.0181	0.0181	0.0181
Ethylene	lbmole/hr	0.0000*	0.0000	0.0000	0.0000
Total:	lbmole/hr	243.1529*	442.7829	442.7829	442.7829

Stream		305	305a	305b	306
Vapour frac.		1.0000	1.0000	1.0000	0.0000
Temperature	F	-134.7006	-139.2325	-80.0000*	-172.8749
Pressure	psia	58.0000	25.0000*	20.0000	659.9999
Enthalpy	Btu/hr	112074.0937	112074.0937	133506.1588	-196405.0151
Methane	lbmole/hr	40.9401	40.9401	40.9401	291.0702
Nitrogen	lbmole/hr	4.1198	4.1198	4.1198	8.7583
Methanol	lbmole/hr	0.0000	0.0000	0.0000	0.0022
CO2	lbmole/hr	0.0001	0.0001	0.0001	0.0122
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	45.0600*	45.0600	45.0600	299.8429

Stream		306a	307	308	308a
Vapour frac.		0.2510	0.0000	0.0000	0.1180
Temperature	F	-223.6690	-172.8749	-223.6690	-253.1212
Pressure	psia	63.0000*	659.9999	63.0000	20.0000*
Enthalpy	Btu/hr	-196405.0151	-93629.4757	-327778.7305	-327778.7305
Methane	lbmole/hr	291.0702	138.7579	222.6972	222.6972
Nitrogen	lbmole/hr	8.7583	4.1752	1.8779	1.8779
Methanol	lbmole/hr	0.0022	0.0010	0.0022	0.0022
CO2	lbmole/hr	0.0122	0.0058	0.0121	0.0121
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	299.8429	142.9400	224.5894	224.5894

Stream		309	310	311	312
Vapour frac.		0.0000	0.0667	1.0000	1.0000
Temperature	F	-253.1212	-184.1126	-134.7006	70.0000*
Pressure	psia	20.0000	200.0000*	195.0000	190.0000
Enthalpy	Btu/hr	-369944.3360	-93629.4757	333515.1658	590767.9029
Methane	lbmole/hr	197.5814	138.7579	138.7579	138.7579
Nitrogen	lbmole/hr	0.4997	4.1752	4.1752	4.1752
Methanol	lbmole/hr	0.0022	0.0010	0.0010	0.0010
CO2	lbmole/hr	0.0121	0.0058	0.0058	0.0058
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	198.0953	142.9400*	142.9400	142.9400

Table D-3, Material Balance for Figure D-3, Continued

Stream		313	314	314a	315
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	-223.6690	-134.7006	-134.7006	70.0000
Pressure	psia	63.0000	58.0000	58.0000	53.0000
Enthalpy	Btu/hr	131381.9518	187172.0244	75097.9307	125968.3110
Methane	lbmole/hr	68.3730	68.3730	27.4329	27.4329
Nitrogen	lbmole/hr	6.8804	6.8804	2.7606	2.7606
Methanol	lbmole/hr	0.0000	0.0000	0.0000	0.0000
CO2	lbmole/hr	0.0001	0.0001	0.0001	0.0001
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	75.2535	75.2535	30.1935	30.1935

Stream		316	317	318	319
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	84.1382	322.9333	100.0000*	-77.9569*
Pressure	psia	53.0000	195.0000	190.0000	680.0000*
Enthalpy	Btu/hr	243934.8159	366768.8771	248020.8800	455830.5402
Methane	lbmole/hr	52.5487	52.5487	52.5487	191.3067
Nitrogen	lbmole/hr	4.1388	4.1388	4.1388	8.3140
Methanol	lbmole/hr	0.0000	0.0000	0.0000	0.0010
CO2	lbmole/hr	0.0001	0.0001	0.0001	0.0059
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	56.6876	56.6876	56.6876	199.6276

Stream		319r	320	321	322
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	-77.9569*	-253.1212	-134.7006	70.0000
Pressure	psia	680.0000*	20.0000	18.0000	16.0000
Enthalpy	Btu/hr	455835.8682	42179.4476	67370.8183	111627.2485
Methane	lbmole/hr	191.3092*	25.1158	25.1158	25.1158
Nitrogen	lbmole/hr	8.3139*	1.3782	1.3782	1.3782
Methanol	lbmole/hr	0.0010*	0.0000	0.0000	0.0000
CO2	lbmole/hr	0.0059*	0.0000	0.0000	0.0000
Ethylene	lbmole/hr	0.0000*	0.0000	0.0000	0.0000
Total:	lbmole/hr	199.6300*	26.4941	26.4941	26.4941

Stream		323	324	325	326
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	300.5362	100.0000*	78.4979	313.6996
Pressure	psia	58.0000	53.0000	190.0000	689.9999
Enthalpy	Btu/hr	167072.1267	117966.4979	838788.7914	1.25478E+06
Methane	lbmole/hr	25.1158	25.1158	191.3067	191.3067
Nitrogen	lbmole/hr	1.3782	1.3782	8.3140	8.3140
Methanol	lbmole/hr	0.0000	0.0000	0.0010	0.0010
CO2	lbmole/hr	0.0000	0.0000	0.0059	0.0059
Ethylene	lbmole/hr	0.0000	0.0000	0.0000	0.0000
Total:	lbmole/hr	26.4941	26.4941	199.6276	199.6276

Table D-3, Material Balance for Figure D-3, Continued

Stream		327	328	LNG
Vapour frac.		1.0000	1.0000	0.0250
Temperature	F	100.0000*	-69.7506	-259.7447
Pressure	psia	685.0000	680.0000	14.6960*
Enthalpy	Btu/hr	829620.8237	477262.5976	-369944.3360
Methane	lbmole/hr	191.3067	191.3067	197.5814
Nitrogen	lbmole/hr	8.3140	8.3140	0.4997
Methanol	lbmole/hr	0.0010	0.0010	0.0022
CO2	lbmole/hr	0.0059	0.0059	0.0121
Ethylene	lbmole/hr	0.0000	0.0000	0.0000
Total:	lbmole/hr	199.6276	199.6276	198.0953

Stream		WK301	WK302	Wk303
Enthalpy	hp	21.7907	48.2756	163.4935

Stream		QHX302	QHX303	QHX303A	QHX303B
Enthalpy	Btu/hr	751421.0634	502482.1676	55790.0726	25191.3701

Stream		QHX303C	QHX304	QHX304A	QHX304B
Enthalpy	Btu/hr	427144.6451	352358.2189	50870.3808	44256.4293

Stream		QHX304C	QHX305	QHX306	Qhx307
Enthalpy	Btu/hr	257252.7310	118747.9956	49105.6272	425166.4920

FIGURE D-4
Ethylene Refrigeration System

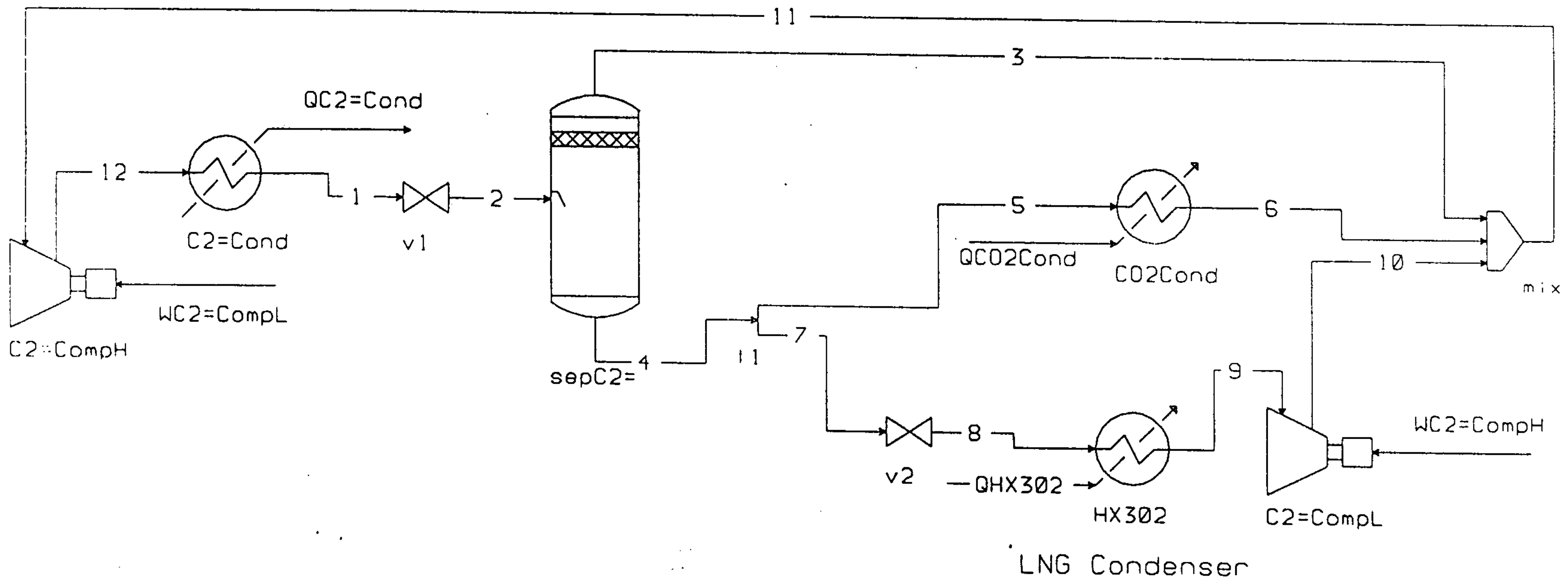

Table D-4, Material Balance for Figure D-4

Stream		1	2	3	4
Vapour frac.		0.0000*	0.2061	1.0000	0.0000
Temperature	F	-30.0000*	-80.0000*	-80.0000	-80.0000
Pressure	psia	250.1904	101.3587	101.3587	101.3587
Molar Flow	lbmole/hr	437.6636	437.6636	90.2174	347.4461
Enthalpy	Btu/hr	-515627.9837	-515627.9837	251827.3870	-767455.3814
Ethylene	mole frac.	1.0000*	1.0000	1.0000	1.0000
Propane	mole frac.	0.0000*	0.0000	0.0000	0.0000

Stream		5	6	7	8
Vapour frac.		0.0000	1.0000*	0.0000	0.1544
Temperature	F	-80.0000	-80.0001	-80.0000	-128.0000*
Pressure	psia	101.3587	101.3587	101.3587	32.5806
Molar Flow	lbmole/hr	188.7840	188.7840	158.6621	158.6621
Enthalpy	Btu/hr	-416995.1114	526959.9110	-350460.2328	-350460.2328
Ethylene	mole frac.	1.0000	1.0000	1.0000	1.0000
Propane	mole frac.	0.0000	0.0000	0.0000	0.0000

Stream		9	10	11	12
Vapour frac.		1.0000*	1.0000	1.0000	1.0000
Temperature	F	-127.9997	-2.3559	-51.7333	62.2812
Pressure	psia	32.5806	101.3587	101.3587	250.1904
Molar Flow	lbmole/hr	158.6621	158.6621	437.6636	437.6636
Enthalpy	Btu/hr	400960.7675	569373.1159	1.34816E+06	1.76393E+06
Ethylene	mole frac.	1.0000	1.0000	1.0000	1.0000
Propane	mole frac.	0.0000	0.0000	0.0000	0.0000

Stream		QCO2Cond	QHX302	QC2=Cond
Enthalpy	Btu/hr	943955.0200*	751421.0042*	2.27956E+06

Stream		WC2=CompH	WC2=CompL
Enthalpy	hp	66.1885	163.4070

FIGURE D-5

Propane Refrigeration System

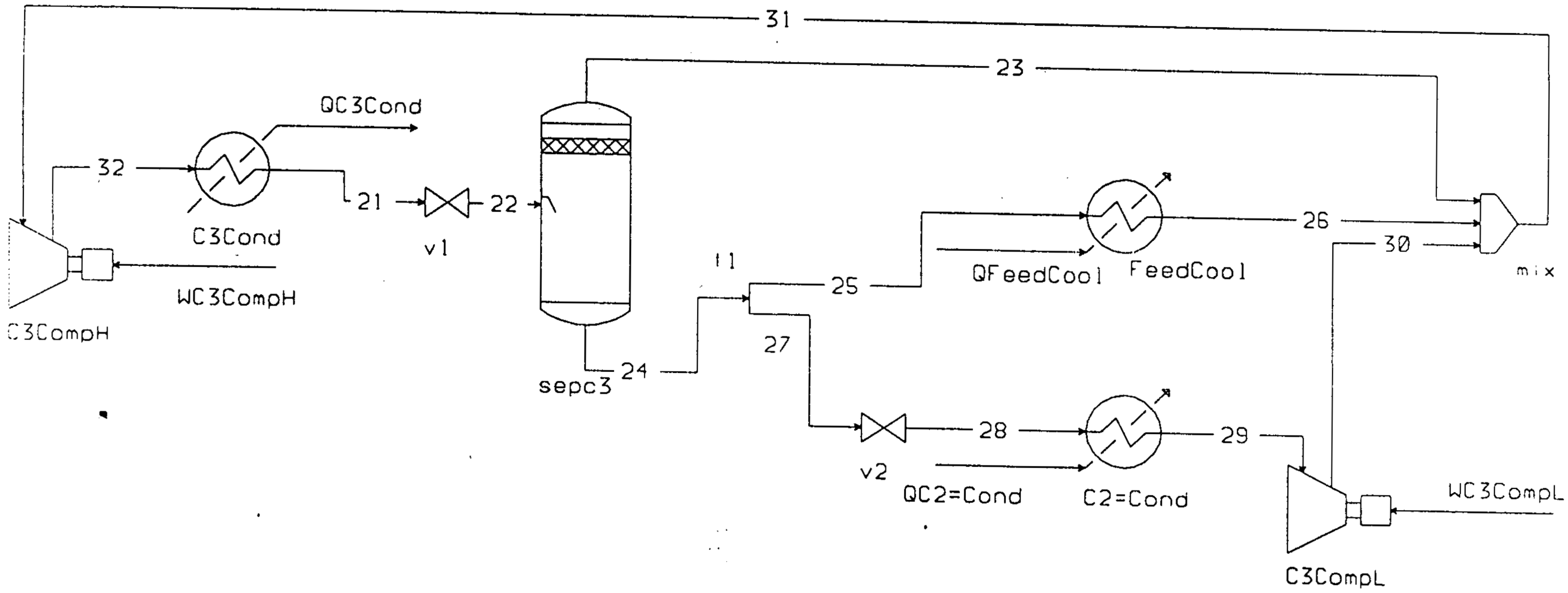

Table D-5, Material Balance for Figure D-5

Stream		21	22	23	24
Vapour frac.		0.0000*	0.2881	1.0000	0.0000
Temperature	F	100.0000*	30.0000*	30.0000	30.0000
Pressure	psia	190.2607	66.6752	66.6752	66.6752
Molar Flow	lbmole/hr	535.6445	535.6445	154.3318	381.3127
Enthalpy	Btu/hr	232732.9199	232732.9199	859036.5927	-626303.6707
Ethylene	mole frac.	0.0000*	0.0000	0.0000	0.0000
Propane	mole frac.	1.0000*	1.0000	1.0000	1.0000

Stream		25	26	27	28
Vapour frac.		0.0000	1.0000*	0.0000	0.2210
Temperature	F	30.0000	30.0002	30.0000	-40.0000*
Pressure	psia	66.6752	66.6752	66.6752	16.0632
Molar Flow	lbmole/hr	19.4300	19.4300	361.8828	361.8828
Enthalpy	Btu/hr	-31913.5696	108150.4342	-594390.0977	-594390.0977
Ethylene	mole frac.	0.0000	0.0000	0.0000	0.0000
Propane	mole frac.	1.0000	1.0000	1.0000	1.0000

Stream		29	30	31	32
Vapour frac.		1.0000*	1.0000	1.0000	1.0000
Temperature	F	-39.9981	70.3274	57.4442	148.1514
Pressure	psia	16.0632	66.6752	66.6752	190.2607
Molar Flow	lbmole/hr	361.8828	361.8828	535.6445	535.6445
Enthalpy	Btu/hr	1.68516E+06	2.27783E+06	3.24501E+06	3.95981E+06
Ethylene	mole frac.	0.0000	0.0000	0.0000	0.0000
Propane	mole frac.	1.0000	1.0000	1.0000	1.0000

Stream		QFeedCool	QC2=Cond	QC3Cond
Enthalpy	Btu/hr	140063.9978*	2.27956E+06*	3.72708E+06

Stream		WC3CompL	WC3CompH
Enthalpy	hp	232.9246	280.9256

FIGURE D-6

Feed Cooler and CO2 Condenser

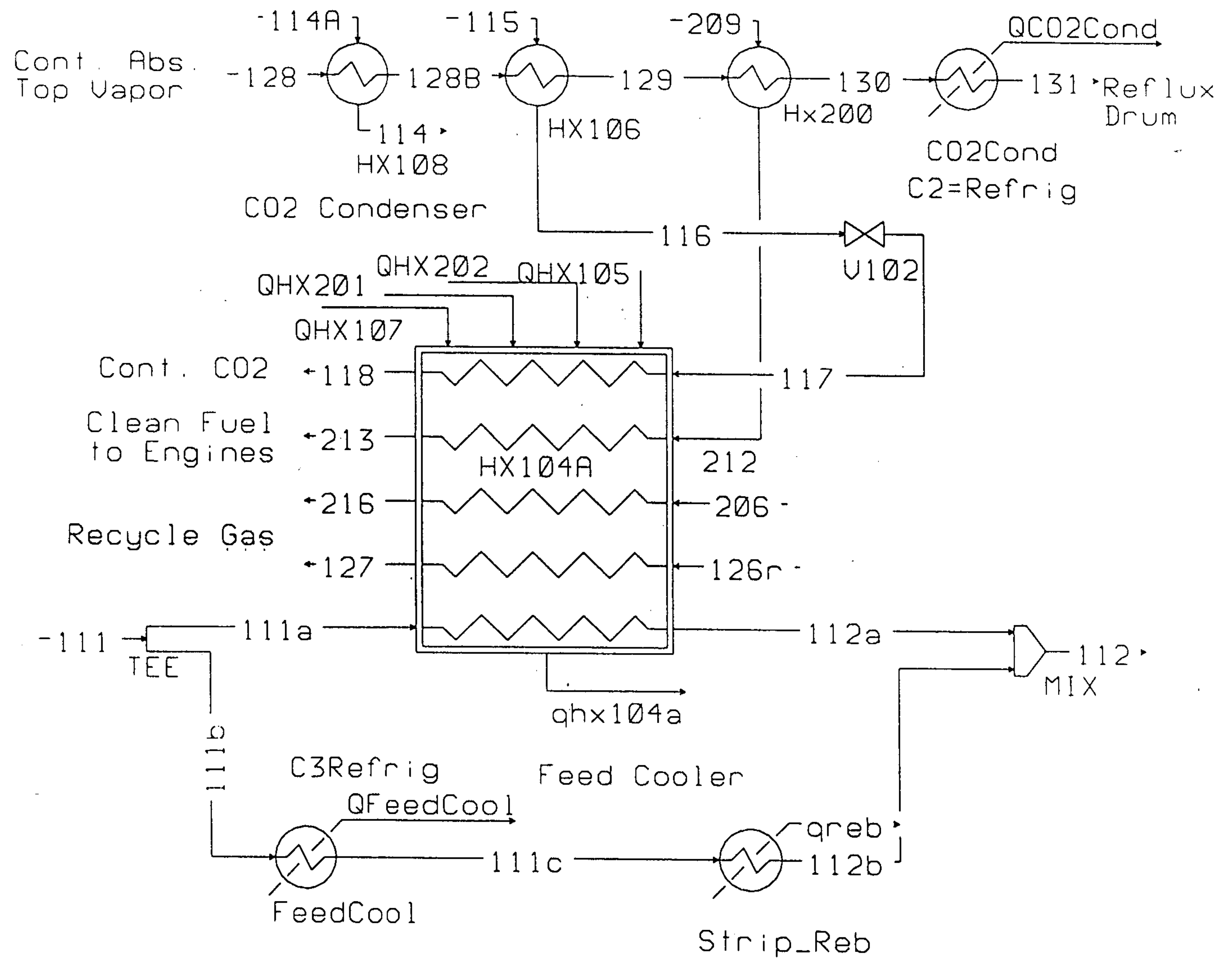

Table D-6, Material Balance for Figure D-6

Stream		111	111a	111b	111c
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	100.0000	100.0000	100.0000	54.5297
Pressure	psia	715.0000	715.0000	715.0000	715.0000
Molar Flow	lbmole/hr	541.7983	264.1800	277.6183	277.6183
Enthalpy	Btu/hr	2.12604E+06	1.03665E+06	1.08938E+06	949325.2135

Stream		112	112a	112b	114
Vapour frac.		1.0000	1.0000	1.0000	0.6763
Temperature	F	7.0000	6.9995	6.9989	-34.0273
Pressure	psia	715.0000	715.0000	715.0000	210.0000
Molar Flow	lbmole/hr	541.7983	264.1800	277.6183	73.5112
Enthalpy	Btu/hr	1.52799E+06	745051.0009	782948.2433	69631.2210

Stream		114A	115	116	117
Vapour frac.		0.2883	0.0974	0.9954	0.9963
Temperature	F	-50.5966	-60.0000	-40.0000	-47.0000
Pressure	psia	210.0000	104.2257	99.2257	69.4984
Molar Flow	lbmole/hr	73.5112	23.7954	23.7954	23.7954
Enthalpy	Btu/hr	-120421.9747	-76409.8851	70334.8883	70334.8883

Stream		118	126r	127	128
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	85.0000	-42.9558	85.0000	-23.5032
Pressure	psia	64.4984	210.0000	210.0000	700.8824
Molar Flow	lbmole/hr	23.7954	121.5323	121.5323	563.8818
Enthalpy	Btu/hr	100901.5808	353300.5000	503682.1366	1.41405E+06

Stream		128B	129	131	130
Vapour frac.		0.9315	0.8801	0.5432	0.8734
Temperature	F	-29.0256	-33.7840	-75.0000	-34.4419
Pressure	psia	700.8824	700.0000	700.0000	700.0000
Molar Flow	lbmole/hr	563.8818	563.8818	563.8818	563.8818
Enthalpy	Btu/hr	1.22400E+06	1.07725E+06	113728.0040	1.05768E+06

Stream		206	209	212	213
Vapour frac.		1.0000	1.0000	1.0000	1.0000
Temperature	F	-64.9602	-81.4964	-44.0000	85.0000
Pressure	psia	20.0000	18.0000	16.0000	15.0000
Molar Flow	lbmole/hr	31.1100	63.3577	63.3577	63.3577
Enthalpy	Btu/hr	92261.8223	184337.6119	203911.1985	273054.5232

Stream		216
Vapour frac.		1.0000
Temperature	F	85.0000
Pressure	psia	15.0000
Molar Flow	lbmole/hr	31.1100
Enthalpy	Btu/hr	133775.7108

Table D-6, Material Balance for Figure D-6, Continued

Stream		qreb	qhx104a	QFeedCool	QHX105
Enthalpy	Btu/hr	166376.9944	291605.5548	140064.0126	150381.6364

Stream		QHX107	QHX201	QCO2Cond	QHX202
Enthalpy	Btu/hr	30566.6933	69143.3243	943955.5532	41513.8878

Table D-7
Equipment Size
LNG +CO2 from 3.5 MMSCFD of Digester Gas
Acrion Contaminant Removal Process
and MeOH Final CO2 Removal

Tag	Service	Quantity	Size		Mat. + Press.
Compressors					
K101	Landfill Gas Stage 1	1	391	hp	
K102	Landfill Gas Stage 2	1	389	hp	
K103	Landfill Gas Stage 3	1	422	hp	
K301	Methane Stage 1	1	22	hp	
K302	Methane Stage 2	1	48	hp	
K303	Methane Stage 3	1	164	hp	
C2=CompL	Ethylene Stage 1	1	66	hp	
C2=CompH	Ethylene Stage 2	1	163	hp	
C3CompL	Propane Stage 1	1	233	hp	
C3CompH	Propane Stage 2	1	281	hp	
			2179	hp	
Exchangers					
HX104	Feed Cooler	1	266	ft^2	SS
FeedCool	Feed Cooler -C3Ref	1	47	ft^2	CS
strip Reboiler	Stripper Reboiler -Feed Cooler	1	64	ft^2	CS
HX106	Contam. CO2 evap.-CO2 Cond	1	59	ft^2	SS
	CO2 Condenser-C2=Ref	1	487	ft^2	Al
HX200	CO2 Cond	1	33	ft^2	Al
HX108	CO2 Cond	1	140	ft^2	SS
	C2=Cond-C3Ref	1	1916	ft^2	Al
HX302	LNG Cond-C2=Ref	1	885	ft^2	Al
HX303	LNG Subcooler	1	259	ft^2	Al
HX304	LNG Recycle Gas Cooler	1	309	ft^2	Al
HX308	LNG Recycle Gas Cooler	1	30	ft^2	Al
	C3 Condenser	1	9318	ft^2	CS
Pumps					
abs100pump	Liquid CO2 Absorbent	2	1	hp	SS
PUMP200	Methanol Solvent	2	9	hp	SS
PUMP201	Methanol Solvent	2	14	hp	SS
Towers			Dia	Ht	
abs100	Contaminent Absorber	1	1.5	45'	CS 700 psi
strip	Methane Stripper	1	1'	30'	KCS 300 psi
ABS200	CO2 Absorber	1	1.5'	40'	SS 700 psi
REG200	CO2 Stripper	1	1.5'	40'	SS 50 psi
Vessels			Dia	Ht	
abs100reflux	CO2 Reflux	1	2.0'	11'	SS 700 psi
sep	CO2 Flash	1	1.5'	5'	KCS 300 psi
SEP201	Methanol Flash	1	2.5'	14'	SS 300 psi
SEP200	Methanol Flash	1	2.5'	9'	SS 20 psi
PS302	LNG Flash 2	1	2.0'	9'	SS 200 psi
PS303	LNG Flash 3	1	2.0'	7'	SS 50 psia
			Dia	Ht	
Dehydration	Mol Sieve Beds+ Heater	2	3.0	12'	CS 300psi
H2S Removal	Sulfatreat Beds	2	5'	15'	CS 300psi
			KGal		
Storage	Liquid CO2 Storage Tank	1	38	KGal	CS 300psi
	LNG Tank	1	66	KGal	SS
Power	Electric Power	1	1136	kW	

CS= Carbon Steel
SS= Stainless Steel
KCS = Killed Carbon Steel

APPENDIX E

Letter Report
Cost Estimate for a Demonstration Unit to Produce Methane from LFG for Liquefaction

Acrion
Technologies, Inc.

June 13, 2000

Dr James Wegrzyn
Brookhaven National Laboratory / Building 815
Upton, New York 11973

Re: DE AC02 99CH10982 / LNG from Landfill Gas, Demonstration Scale

Dear Jim,

Acrion has completed its preliminary look at a demonstration landfill gas processing plant to make methane for liquefaction. A process design for 1200 gallons per day, including process instrumentation, was developed and submitted to recognized vendors for quotes. The design does not include liquefaction. The quotes are budgetary and believed in the range ±15%.

Process Unit	Function	Vendor	Cost
Compression	0 to 425 psig	J-W	$150,000
CO2 Wash	Remove VOC, bulk CO2	Wittemann	$160,000
Membrane	Remove CO2 to 1%	Cynara ($250,000) UOP ($170,000)	$170,000
Mol Sieve (TSA)	Remove CO2 to 50 ppm	UOP	$330,000
		Budget Estimate:	$810,000

Process units are skid mounted, a compression skid and process skids. Installation, which involves utilities, footings and interconnects, is estimated at $25,000 per skid. Hence, Acrion's first pass cost estimate is $860,000 (±15%).

Vendor price quotes generally agree with Acrion's internal estimates except for trace CO2 removal, which is about three times higher than expected. This surprise may be due to several factors: 1) UOP's experience lies with large chemical plants, not small modular skids; 2) UOP emphasized low operating cost rather than low capital cost; 3) UOP used proprietary molecular sieve rather than lower cost generic adsorbents; and 4) UOP is probably telling us politely the job is too small.

Acrion is pursuing the task of firming the cost estimate and identifying lower cost technologies and sources for trace CO2 removal. We believe trace CO2 removal cost can be reduced by a factor of two to three and, coupled with small improvements in the other process units, will yield an installed package price of about $600,000. A second alternative to reduce cost is fabrication of the entire demo unit by one vendor, raw landfill gas to liquid methane in storage; Acrion has identified a one-stop vendor, and will purse this option with your approval.

Your guidance to define demonstration objectives and establish cost targets would help us negotiate with vendors and expend efforts where they are needed most. Please call to discuss at your convenience, 216-573-1187.

Sincerely,

Bill

William R Brown
President

cc: Jeff Cook / Larry Siwajek

9099 Bank Street, Cleveland, Ohio 44125
(216) 573 1187
FAX (216) 573 1186

Acrion
Technologies, Inc.

October 20, 2000

Dr Wai Lin Litzke
Brookhaven National Laboratory
Building 815
Upton, New York 11973

Re: Phase II Final Report, Contract 725089

Dear Wai Lin

Enclosed are two copies of the Phase II Final Report on Landfill Gas Conversion to LNG and LCO2 for you and Jim Wegrzyn. I have also sent a copy to Patrice Brewington at Argone in Chicago.

Acrion is interested in a commercial or demonstration scale LNG project at Waste Management's Arden Landfill in Washington PA as a Phase III effort. We would like to define a Phase III effort with you.

Our LFG CO2 recovery demonstration skid has just been completed and we will begin installation at the New Jersey Ecocomplex near Trenton NJ in November. When it is operational we would like to have you and Jim come and see it.

Sincerely,

Jeff

Jeff Cook

enclosure

9099 Bank Street, Cleveland, Ohio 44125
(216) 573 1187
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