SYSTEMS ANALYSIS OF MUNICIPAL SOLID WASTE BIOGASIFICATION

Technical Report

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NOTICE

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PREFACE

This report describes the work carried out on the project "Systems Analysis of Municipal Solid Waste Biogasification," by Hunter/RS&H* for the Solar Energy Research Institute (SERI). This work has consisted of systems analysis and engineering support to research conducted at the Walt Disney World Resort Complex near Orlando, Florida. This research was initiated by Walt Disney Imagineering and the Gas Research Institute (GRI) in the early 1980s. Originally, it was focused on sewage treatment with aquatic plants, but eventually (by 1986) came to encompass integrated nonenergy intensive treatment of a wide variety of community-derived wastes. Other participants include the Bioprocess Engineering Research Laboratory at the University of Florida (UF), the Institute of Gas Technology (IGT) in Chicago, and the University of Illinois.

We would like to acknowledge the help of V.J. Srivastava (IGT); B.T. Goodman, P.W. Bergeron, N.D. Hinman, C.J. Rivard, and K. Grohmann (SERI); D.P. Chynoweth and T. Chen (UF); T.D. Hayes (GRI); and C.A. Stokes (Stokes Consulting Group--methanol) for the pertinent information they provided. S.M. Skinner, E. Crews, K. Berry, and N.A. Kiger patiently and efficiently typed the report and saw it through many revisions. Thanks are due also to W.T. Todd for thoroughly reviewing the draft, and to Dr. P.F. Hutchins for reviewing Appendix B.

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SUMMARY

OBJECTIVES

The primary objective of this project was to develop a mathematical, computer-based model of biogasification of municipal solid waste (MSW) to be used to focus future research in the DOE/SERI Anaerobic Digestion Program. This objective was addressed through the following tasks: (1) information gathering, (2) development of biogasification model, (3) sensitivity analyses, and (4) research recommendations.

DISCUSSION

A spreadsheet-based computer model of biogasification of MSW was designed. It includes modules for preprocessing of refuse; feed input; conversion of feed through anaerobic digestion; gas utilization options of medium Btu gas, synthetic natural gas (SNG), gas turbine, and methanol; residue post-processing options of landfilling, incineration, power generation, and composting; energy balance; cost estimation; financial assumptions and levelized cost-of-service calculations. The model calculates the levelized cost of the energy product (medium Btu gas, SNG, electricity or methanol) produced by the facility broken down into the costs attributable to the different modules. It also calculates the revenues from tipping fees and the sale of recyclable materials. A series of sensitivity analyses were performed using the model. Parameters which were evaluated over a range of values or options included: facility size, facility availability, disposition of plastics, conversion technology, solids concentration of digesters, retention time in digesters, kinetic reaction rate, residue post-processing options, gas utilization options, tipping fee, value of recyclables, and electricity cost and value. Based on the sensitivity analyses, cost-optimal cases for each conversion technology were developed. Research recommendations were made based on these optimal cases.

CONCLUSIONS

1. The MSWAD (Municipal Solid Waste Anaerobic Digestion) computer model can be used to compare technologies and operational parameters, explore combinations of subprocesses, enter various costs, rates, chemical characteristics, etc., to allow optimization based on local conditions. It can be used to measure the sensitivity of cost or energy efficiency to technological choices and operational parameters.

2. Retention time should be approached as a design variable to be optimized after all other parameter values have been established. The biogasification process is not very sensitive to retention time and economically optimal values are on the order of a few weeks.

3. For this base process combination, economies of scale are observed up to 500 tpd. The economies would probably extend to larger sizes for a more capital-intensive process (e.g., with combustion of solid residues).

4. Reaction rates have a significant impact on cost up to first-order rate constants of approximately 0.25 day^{-1}.

5. Methane Enrichment Digestion (MED) has the potential of reducing costs by \$0.95/MMBtu, but a more thorough evaluation is necessary. MED does not require an absorber, which reduces capital cost compared to conventional absorption technology, and only the product SNG needs to be compressed, which reduces the operating cost.

6. The most economical way to deal with the solid residue is to refine it to compost and market it as such, because this diverts residue from the landfill and minimizes landfilling expenses. However, the continued marketing of hundreds of tons of compost per day is a major management challenge. A considerable amount of regulatory uncertainty and questions about present and future environmental impacts of MSW-derived compost application remains.

7. Marketing medium Btu (500-600 Btu/scf) gas is the most economical biogas utilization option but also the most difficult to arrange since it requires a long-term contract with a nearby large user.

8. Upgrading the biogas to pipeline quality gas (SNG) is more expensive but results in a more marketable product. MED has the potential to considerably narrow the gap between medium and high Btu gas production.

9. The relationship between the tipping fee charged by an MSW anaerobic digestion facility and the price at which it has to sell its energy product is inverse and linear.

10. If the process train is designed for maximum energy recovery (combustion of all solid residues with power generation and marketing of medium Btu gas), the three processes recover the following percentage of the MSW's energy content: RefCoM = 46 percent, SOLCON = 70 percent, SERI HS = 64 percent.

11. The most energy efficient process is not necessarily the most economical.

12. If energy products are priced competitively (\$3/MMBtu for gas, 3¢/kWh for electricity, 50¢/gallon for methanol), the following breakeven tipping fees result, for optimized processes. RefCoM: \$32-37/ton MSW; SOLCON: \$23-30/ton; SERI HS: \$20-24/ton.

13. Of the three processes considered, REFCOM IS THE ONLY TECHNOLOGY THAT HAS INDUSTRIAL CREDIBILITY, having been demonstrated at 20 tpd. SOLCON has only been operated at pilot scale (0.05 tpd) and SERI HS at bench scale (0.001 tpd).

14. Anaerobic digestion of MSW is cheaper and cleaner than burning or landfilling, and it allows significant energy recovery in the form of fuel (including methanol), produced year-round at the point of usage, namely the city. Anaerobic digestion is an economically superior source of fuel because it is subsidized by the tipping fees. This subsidy reflects society's need for waste disposal and is therefore real and permanent. Anaerobic digestion requires a substantially lower tipping fee than established solid waste technologies and is therefore economically superior to them.

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NOMENCLATURE

BMP	Biochemical Methane Potential
BVS	Biodegradable Volatile Solids
C_A	Concentration of Compound A (Moles L^{-1})
COD	Chemical Oxygen Demand, a measure of total organics concentration, typically used for characterization of wastewater
COWSA	Community Waste Systems Analysis, an earlier computer model
CSTR	Continuously Stirred Tank Reactor
DOE	U.S. Department of Energy
DRANCO	Dry Anaerobic Composting, an MSW biogasification process
ETU	Experimental Test Unit
GRI	Gas Research Institute
HHV	Higher Heating Value
HRT	Hydraulic Retention Time
IGT	Institute of Gas Technology
k	First order reaction rate coefficient
L	Liter
MBG	Medium Btu Gas (400-600 Btu/scf)
MED	Methane enrichment digestion
MGD, mgd	Million Gallons per Day
MMBtu	Million British Thermal Units
MSW	Municipal Solid Waste
MSWAD	Municipal Solid Waste Anaerobic Digestion, a computer model
NMVFR	Non-Mixed Vertical Flow Reactor
O&M	Operation and Maintenance
PDA	Process Development Allowance
r_A	Reaction Rate (moles L^{-1} s^{-1})
rpm	Revolutions per minute
RDF	Refuse-derived feed
RefCoM	Refuse Converted to Methane, an MSW biogasification process
RS&H	Reynolds, Smith and Hills, A-E-P, Inc. (now Hunter/RS&H)
s	Second
scf	Standard Cubic Foot, at 1 atmosphere (14.7 psi) and 0°C (32°F)
SERI	Solar Energy Research Institute
SERI HS	SERI high solids digester
SNG	Synthetic Natural Gas (>950 Btu/scf)
SOLCON	Solids Concentrating digester, an MSW biogasification process
SRT	Solids Retention Time
STP	Standard Temperature and Pressure
t	Space Time (= hydraulic retention time)
TPD, tpd	U.S. Tons Per Day
TS	Total Solids, or dry solids remaining after removal of moisture
UF	University of Florida
VS	Volatile Solids, or combustible solids
VSCE	Volatile Solids Conversion Efficiency
v/v/d	Volumes (of gas) per unit reactor volume per day

Subscripts:

a	Added
A	Compound A

CONVERSION TABLE

Btu	= British thermal unit	= 1,055 Joules
Btu/lb	= Btu per pound	= 2,326 Joule/kg
cf, cu ft	= cubic foot	= 28.32 liter
gal	= gallon	= 3.7854 liter
g VS L^{-1}	= grams VS per liter	= 0.0624 lb/cu ft
hp	= horsepower	= 0.7457 kilowatt (kW)
in	= inch	= 2.54 cm
L	= liter	= 0.0353 cu ft
lb	= pound	= 0.4536 kg
lb/cu ft	= pound per cubic foot	= 16.019 g L^{-1}
mgd	= million gallons per day	= 3,785 m^3/day
ml/g	= L/kg	= 0.01602 cu ft/lb
MMBtu	= Million Btu	= 1.055 GJ (Gigajoule, 10E9 Joule)
quad	= quadrillion Btu	= 10E15 Btu = 1.055EJ (Exajoule, 10E18 Joule)
scf	= standard cubic foot	= 28.32 liter, at 1 atmosphere (101.33 kPa) and 0°C
scf/lb VS_a	= standard cubic foot (of methane or biogas) per pound volatile solids added to the digester	= 0.0624 liter per gram volatile solids added
ton	= customary U.S. ton	= 2,000 pounds = 0.9072 Mg (megagram, metric ton)
tpd	= U.S. tons per day	= 0.9072 Mg/day
95°F	= 35°C	
131°F	= 55°C	
$	= 1990 U.S. dollars	

1.0 INTRODUCTION

This report describes the development and use of a systems analysis model for municipal solid waste. A narrative description of the model is provided in Section 2. This model is named MSWAD (Municipal Solid Waste Anaerobic Digestion) and was developed from an earlier model referred to as COWSA, which is described in the 1988 Annual Report submitted by Hunter/RS&H to SERI. The differences between the two models are listed in Section 2 of the present report.

Once the MSWAD model was functioning, a base case was developed as described in Tables 1 and 2. The most proven MSW biogasification technology, RefCoM, was selected for this base case, and some conservative assumptions were made regarding process configuration: all residues are landfilled and the biogas is upgraded to SNG. Starting from this base case, a series of sensitivity analyses were carried out by varying one parameter at a time. This is described in Section 3.0 through 3.2 and the parameters analyzed are listed in Table 3.

Next, three different biogasification technologies (RefCoM, SOLCON and SERI High Solids) were considered and compared as to cost and energy efficiency in Sections 3.3 and 3.4. Finally, the potential of these three technologies was explored by considering optimal process configuration in Section 3.5. The report is concluded with some research recommendations in Section 4.

2.0 MODEL DEVELOPMENT

A mathematical model of a community waste biogasification facility was developed; it is known as the MSWAD (Municipal Solid Waste Anaerobic Digestion) model. The model calculates mass balances, energy balances and levelized cost of service. The relationship between process modules and options are shown in Figure 1. The conversion facility accepts MSW, industrial waste and any other appropriate feedstock. MSW must first be processed including shredding, magnetic separation, sorting, etc.; this is quantified in the Preprocessing Module: the quantity and quality of raw trash is entered here and, among others, the amount of refined digester feed is calculated. The generated feedstock information, plus data about any other feedstocks that may be considered, is entered into the Feed Input Module where the data are processed for further use in the conversion module.

The following residue processing options are available here:

1) Landfilling;
2) Incineration without power recovery;
3) Burning in a solid fuel boiler with full power generation;
4) Marketing as compost.

The biogasification process generates biogas, a medium Btu gas (400 to 600 Btu/scf), which can be utilized in the following ways:

1) Sale of medium Btu gas;
2) Cleanup to SNG(950 Btu/scf);
3) Generate electricity for sale using a gas turbine;
4) Produce methanol for sale.

Information about energy usage and production in every step is processed in the Process Energy Module. Finally, capital costs, O&M, energy product income, etc., are used in the Cost Module to generate a levelized cost of

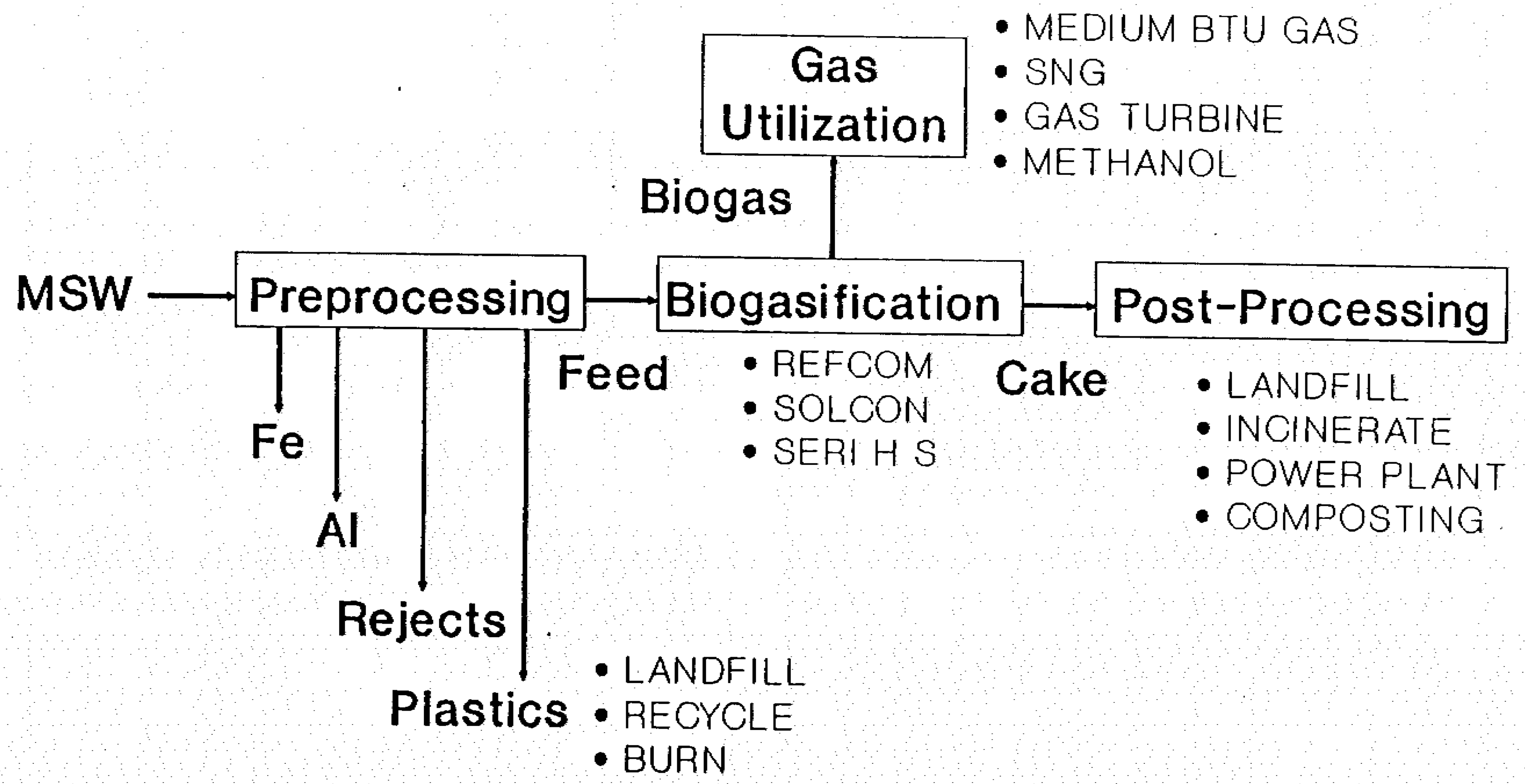


Figure 1: Process modules and options: three options are available to deal with a plastics-enriched stream separated up front; three digester designs were considered; four gas utilization options are included; and four options were investigated to dispose of the solid residue.

service. The remainder of this chapter describes in some detail the major inputs, calculations, and outputs of the process modules. Actual printouts of the modules in the spreadsheet are shown in Appendix A. The basecase assumed for these printouts is summarized on Table 1. For additional details concerning the model, refer to the 1988 Annual Report submitted by Hunter/RS&H to SERI in December 1988.

The following modifications were made to the community waste model (COWSA) described in the 1988 Hunter Annual Report to produce the MSWAD model.

1. Sewage treatment module was removed.
2. Wet oxidation section was removed.
3. Cost of service is expressed in $/MMBtu of energy product output rather than $/ton MSW input.
4. A comprehensive thermal energy balance flowsheet was added.
5. Updated costs were obtained for high solids pumping, gas and electric interconnection, and composting.
6. Medium-Btu gas sale was included as a biogas processing option.
7. Sale of electricity generated by biogas-fired gas turbines was included as a biogas processing option.
8. Conversion of biogas to methanol was included as a biogas processing option.
9. The Methane Enriched Digestion (MED) process was simulated.
10. Landfill life extension calculations were added.
11. The entire spreadsheet was reorganized and streamlined to ensure that all input sections would be together.

In the model, system boundaries coincide with facility boundaries; no collection costs are included nor are societal cost benefits accounted for.

Table 1. Summary of Base Case Assumptions

- Community waste facility treats 500 U.S. tons per day (tpd) of municipal solid waste (MSW).
- The 1974 EPA analysis of MSW was used to estimate TS, VS and BVS.
  - 74% Total Solids (TS)
  - 73% Volatile Solids (VS) of TS
  - 87% Biodegradable Volatile Solids (BVS) of VS
  - 0.45% Aluminum
- The RDF produced by preprocessing has the following characteristics:
  - 443 tpd RDF
  - 62.3% TS
  - 86.4% VS (of TS)
  - 91.9% BVS (of VS)
- Plastics-enriched stream from preprocessing is landfilled.
- The RefCoM digester technology is used, consisting of continuously stirred tank reactors (CSTR) operating at the following conditions:
  - Hydraulic Retention Time (HRT) = 23 days
  - Loading Rate = 0.37 lb VS/cf-day
  - Methane Yield = 5.0 scf methane/lb VS added
  - Methane Production Rate = 1.84 vol. methane/vol. reactor/day
  - Reactor Solids Content = 8.0% TS
- Biogas is cleaned up to SNG and sold.
- The undigested residue is dewatered and landfilled.
- Economic Assumptions:
  - Facility tipping fee = landfill tipping fee and hauling = $40/ton
  - Cost of electricity = $0.05/kwh
  - Costs are expressed in 1990 $
  - Financing cost = 40% of capital cost

2.1 PRIMARY INPUT SECTION

The model user enters values which describe the overall plant configuration in the Input Section. These inputs include MSW quality, quantity, and tipping fee; selection of solid residue processing option; selection of gas utilization option; and selection of base year for levelized cost (in other words, in what year's dollars should the cost be expressed). The user also enters values for conversion variables such as solids retention time (SRT), ratio of solids retention time to hydraulic retention time (SRT/HRT), digester temperature, and percentage methane in the biogas produced. These inputs are used throughout the process modules.

2.2 MASS/ENERGY BALANCE ASSUMPTIONS SECTION

The assumptions section is an area in the spreadsheet where the user can specify certain operational parameters to be used in the remaining modules - Preprocessing, Conversion, Biogas Processing, Solid Residue Processing, and Process Energy. Inputs for conversion include desired TS% at various stages in the conversion process, and digester size and quantity limitations. Other inputs concern weather and operating conditions.

2.3 PREPROCESSING MODULE

The Preprocessing Module accepts MSW as the input, which passes through a shredder for size reduction, magnets for removal of ferrous metals, disc screens for size classification, an air stoner and air knife for separation of light and heavy materials, an aluminum recovery system, and a plastic/paper separator. The plastics are separated from paper in a wet trommel, which results in a plastics-enriched stream containing roughly half of the plastics in the MSW and a paper-enriched stream which is suitable for digester feed. The output from the module shows the quantities and qualities of recovered ferrous metals, aluminum, and plastics. The refuse-derived feed (RDF) which results from the separation process is described in some detail, including moisture content, biodegradable and non-biodegradable organics, and inorganics. This constitutes the primary feed to the digesters. The plastics-enriched stream will proceed to one of three treatments: landfill, recycle (at no value or cost), or burn (if the incineration or power plant solid residue processing option is chosen). The remainder of the material, largely mineral and unrecyclable residue, can be landfilled directly. The information about the RDF stream is then automatically transferred to the Feed Input Module.

It must be noted that the Preprocessing Module can be bypassed, for a case where RDF is immediately available. In this case, a negligible MSW inflow can be entered in the Primary Input Section, and the actual RDF amount and quality can be entered directly in the feed input section, as a secondary feedstock (see below).

2.4 FEED INPUT MODULE

The RDF generated in the Preprocessing Module is placed automatically in the appropriate slot of the Feed Input Module. The purpose of this module is to allow input of other feeds to the biogasification systems, such as industrial or agricultural waste. The quantity of feed is entered and parameters for each stream include percent total solids (TS%), percent volatile solids (VS%), and estimates of biodegradability (BVS/VS%), first order reaction rate coefficient (k), and methane yield (COD/VS).

To eliminate the confusion between methane yield per unit of substrate fed and per unit converted, the COD/VS mass ratio (grams COD/gram VS) is used to describe the chemical makeup of the feed and its impact on methane production. One gram of COD yields approximately 350 ml of methane @ STP upon full conversion (5.61 scf methane/lb COD converted). So a substrate with a

COD/VS ratio of 1.2 will yield 5.61 X 1.2 = 6.73 scf methane per pound of volatile solids converted. If the volatile solids conversion efficiency is 75%, then the methane yield will be 6.73 X 0.75 = 5.05 scf methane per pound of volatile solids added to the digester.

The other purpose of the Feed Input Module is to calculate the average composition of the mixed feed, and to estimate the resulting gross biogas production. The streams are summed by the module, and quantities and qualities of the mixed feed stream are calculated. Weighted average volatile solids conversion efficiency (VSCE) and methane production are some of the outputs of this module which are subsequently used in the Conversion Module.

2.5 CONVERSION MODULE

2.5.1 General Comments

RDF will generally be the primary feed to the Conversion Module. Prior to entry into the digester, the feed is blended with recycled filtrate to achieve a higher moisture content. Biogas and methane production from anaerobic digestion are calculated, as well as size and number of digesters. Digesters may be either continuously stirred tank reactors (CSTR) or non-mixed vertical flow reactors (NMVFR). The MSWAD model will calculate the power required for mixing a low solids CSTR. Dewatering of the effluent yields filtercake, which proceeds to a Solid Residue Processing option, and filtrate (liquid). The filtrate is recycled in order to conserve moisture, heat, nutrients, and alkalinity. The module calculates the quantity of excess filtrate to be disposed of or, if necessary, the additional dilution water required.

2.5.2 Reactor Volume and Solids Concentration

As long as a digester is flooded, i.e., is filled with a liquid with suspended solids in it, the volume will be defined by the solids concentration in the liquid. However, when the specified solids concentration yields more dry matter per unit volume than the bulk density allows, the bulk density becomes limiting and defines the volume. In this case, the digester is unflooded, i.e., the liquid level is below the level of the solid ,‑le of biomass, or no separate liquid phase may be present at all. Example: Assume that a particular batch of RDF has a bulk density of 150 dry grams per liter; if the digester is operated at 8% TS, and using the simplifying assumption that the densities of water and solid are identical (approximately 1,000 g/l), this translates to operating the digester with a slurry density of 80 dry grams per liter. In this instance, the digester is flooded and the volume calculation is derived from the mass balance. If the digester is operated at 30% TS, the assumption of a flooded digester would yield roughly 300 dry grams per liter, which is clearly impossible, since only 150 dry grams of RDF can be packed into one liter. In this case, the bulk density is limiting and is used to calculate the volume. For further information on the kinetics and mass balance calculations in the Conversion Module, refer to the 1988 Annual Report.

2.5.3 Reaction Kinetics

The different types of digesters considered were modeled as continuously fed, continuously mixed reactors (CSTR). The conversion reaction was approximated as a first order process ($-r_A = kC_A$, k = reaction rate constant) converting the biodegradable volatile solids (BVS) to gas. It was further assumed that the process occurs at constant volume since it usually operates at a high moisture content. The extent of conversion for a constant volume first order reaction in a CSTR is: $kt/(1 + kt)$ with t = space time (retention time) for the reactor. In this model, the solids retention time (SRT) is used as the space time. How extent of conversion is related to time and reaction rate is illustrated in Figure 2.

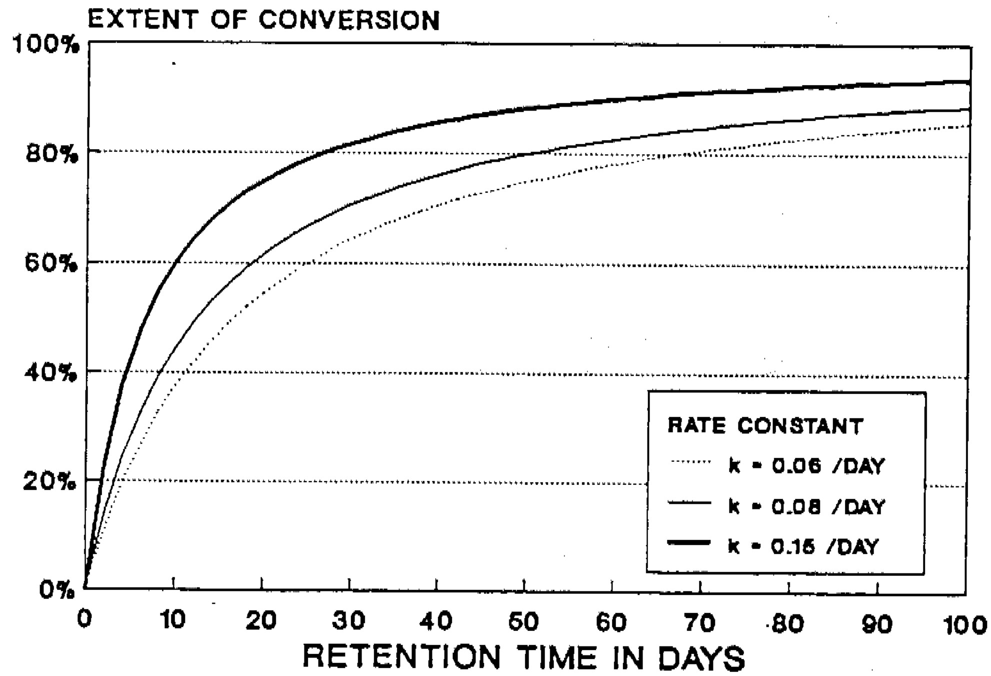


Figure 2: Conversion of biodegradable material versus time, for three first order reaction rate constants k. The rate constants are expressed in day^{-1}. First order chemical reaction kinetics ($-r_A = kC_A$) and a completely mixed continuously fed regime are assumed. Base case: $k = 0.155$ day^{-1}, solids retention time = 23 days.

The constant volume assumption breaks down at the very high solids concentrations as investigated by SERI, since in that case the reacting material significantly diminishes in volume. How this was dealt with is further explained in Section 3.3, Digester Technology Cost Sensitivity.

2.5.4 Methane Enrichment Digestion (MED)

A process to produce high Btu gas directly from anaerobic digesters was invented at GRI and analyzed in the MSWAD model. It is referred to as Methane Enrichment Digestion (MED). The liquid phase inside the digester is used as a CO_2 carrier; this liquid is continuously circulated through a stripper column where CO_2 is removed by a counter current flow of air. The CO_2-depleted liquid is then reintroduced to the digester where it can absorb more CO_2. The process has the advantage over conventional absorption in that it does not require a separate absorber: in effect, the digester itself functions as the absorber.

MED was theoretically developed and subsequently modeled at GRI. It was demonstrated at laboratory scale at the University of Illinois and at Cornell University, and was successfully demonstrated at pilot scale at Walt Disney World.

The GRI computer model written by T.D. Hayes was incorporated in the MSWAD model. The intent was to transfer key data from the MED model to the MSWAD model. Due to basic differences in the design of both models, this effort was only partially successful. Finally, it was decided to use the operating parameters of the successful Walt Disney World tests as the basis for estimating the cost of MED. These parameters are:

- liquid recirculation rate = 1.5 liquid volumes/reactor volume/day,
- air-to-water ratio in stripper = 30:1,
- ratio of digester volume to stripper volume = 20:1.

Furthermore, it was assumed that the air blower only had to overcome a head of 5.5 inches of water due to the very open design of plate tower used for the stripping off CO_2. This open design is necessary to prevent clogging because the carrier liquid can contain a considerable amount of suspended solids. Should the head be higher, the power costs could become prohibitive.

Methane loss was assumed to be equal to that used to model the membrane gas purification process (SNG option), namely ten percent. This was done to ensure an equal energy production on both sides and thus allow a direct comparison of costs, since costs are expressed per unit of gas produced. Curves were developed for the stripping tower, pump, and blower costs, and pump and blower power consumption.

2.6 BIOGAS PROCESSING MODULES

The biogas produced by anaerobic digestion in the conversion process is treated in the biogas processing modules using one of four options: (1) medium Btu gas sale, (2) gas cleanup to SNG, (3) sale of electricity generated by biogas-fired gas turbines, or (4) sale of methanol produced from biogas. The medium Btu gas sale option involves low-pressure compression of biogas to transport it over short distances to an end user.

2.6.1 Gas Cleanup to SNG

The Prism Separator gas cleanup system is simulated using the data provided by Permea, Inc. High pressure biogas passes through the membranes, resulting in an SNG stream containing 95% methane and a low pressure permeate sidestream which is primarily CO_2, but which also contains some methane. Some of this sidestream is repressurized and recycled to boost methane recovery. It is assumed that 90% of the methane is recovered as SNG since higher methane recoveries are associated with rapidly escalating costs. The permeate is fired by the boiler in the incineration with steam and electric generation option and is flared if no burning is included in the system.

2.6.2 Gas Turbine

The biogas stream is fired in a gas turbine to produce electricity. The energy output is calculated by the following equation:

$$\text{KW output} = 43.06\ (\text{MMBtu/hr input})^{1.1588}$$

This equation results in better heat rates for larger gas turbines; for the base case, the heat rate is 11,200 Btu/kwh (30 percent energy efficiency). Heat is recovered from the gas turbine and is used to satisfy process heat requirements, if necessary. The power produced is used for process electricity requirements, with the remainder being sold to the grid. When the gas turbine option is chosen, the levelized cost of service for the facility is expressed in terms of \$/MMBtu (or \$/kwh) of electricity for the facility to break even.

2.6.3 Methanol

The ideal feedstock for methanol manufacture is a mixture of 75 percent CH_4 and 25 percent CO_2. Biogas at 55 to 60 percent CH_4 is close to this ratio and is thus a better feedstock than natural gas. The latter is presently used for methanol manufacture. Some CO_2 will normally be removed from the biogas to bring the methane content up to 75 percent. This can be done using absorption or membrane technology, or using the MED process (see Section 2.5.4). The alternative to CO_2 removal is to buy natural gas to boost the biogas-natural gas mixture's methane content to 75 percent. With biogas containing 55 percent CH_4 (our basic assumption), this would be prohibitively expensive, so it was not considered here.

To manufacture methanol, the 75/25 CH_4/CO_2 mixture is first steam reformed to synthesis gas containing 33 percent CO and 67 percent H_2; subsequently, this synthesis gas is catalytically converted to methanol. Methanol manufacture from natural gas is a mature technology with plants is operation in the U.S.

at different scales. After surveying the state of the art, the following key assumptions were used:

- The capital cost is \$1.0/gallon per year of methanol produced, irrespective of size. Larger plants benefit from economies of scale but these are counterbalanced by increased investment for energy recovery. Energy recovery equipment is unaffordable at the smaller scales.
- The energy efficiency of larger plants is higher. A 25-ton-methanol-per-day plant will require 110 scf of methane per gallon of methanol (50 percent energy efficiency of product versus raw material). A 250-ton-per-day methanol plant requires only 90 scf methane per gallon methanol (61 percent energy efficiency).

In the methanol module of MSWAD, it is assumed that biogas is cleaned up to 75 percent methane using the membrane system described in Section 2.6.1, then converted to methanol using a package methanol plant. The levelized cost of service for the facility is expressed in \$/MMBtu of methanol (or \$/gal) for the facility to break even.

2.7 CONVERSION/BIOGAS PROCESSING OUTPUT SUMMARY

Key data concerning conversion and biogas processing are summarized in this section of the spreadsheet. This summary serves as a data transfer station for subsequent energy and cost calculations. The Solid Residue Processing Modules, Process Energy Module, and Cost Module use the values in this area for calculations.

2.8 SOLID RESIDUE PROCESSING MODULES

Four options have been identified for disposal of the filtercake from conversion. These options are:

1. Landfilling;

2. Incineration with steam generation to meet process heat requirements;
3. Incineration with steam and electricity generation to meet process heat and electric requirements;
4. Composting.

2.8.1 Landfilling

This option may be a viable alternative since a volume decrease of approximately 75% is achieved through anaerobic digestion. Furthermore, the dewatered residue is biologically stabilized which should minimize problems with odor, landfill gas, and organic acid leaching.

2.8.2 Incineration with Steam Generation

The mechanically dewatered filtercake is dried thermally using hot exhaust gases from the boiler and then combined with the plastics from the refuse separation process, if desired, to yield the boiler fuel. The module calculates the amount of water to be evaporated in thermal drying such that the boiler fuel higher heating value (HHV) is at least 4500 Btu/lb, a level which assures good burning and compliance with dioxin regulations mandating 1800°F for at least one second. The boiler is sized to meet, but not exceed, design process heat requirements. A scrubber and precipitator are also accounted for in the mass balance. The ash from combustion falls in a quenching pit, is dewatered and landfilled with the scrubber additive and any precipitate formed during scrubbing.

2.8.3 Incineration with Steam and Electricity Generation

In this option, the filtercake from the digestion process is thermally dried and combined with the plastics, if desired, yielding a boiler fuel with a HHV of at least 4500 Btu/lb. Waste permeate gas from the gas cleanup process is also fired in the boiler; about ten percent of the gross methane production would otherwise be lost with this permeate gas. Its methane concentration is too low for independent combustion. The boiler is sized to obtain the maximum possible steam output. After process heat requirements are met, the remaining steam enters the turbine/generator. The electricity produced is applied to meet process electric requirements. If excess electricity is generated, it is sold as a byproduct at a specified rate and the plant operates as a cogenerator. If additional electricity is required, it is purchased at another specified rate for supplemental power. Here again, the ash is landfilled and state-of-the-art air pollution control equipment is included.

2.8.4 Composting

Compost is a fibrous organic product, usually prepared through aerobic degradation of plant biomass, and used mainly to improve the physical properties of the soil. In heavy soils, it will improve aeration by forming aggregates; in sandy soils, compost will improve the water holding capacity. As a secondary benefit, it can provide slowly released plant nutrients such as nitrogen, phosphorus, potassium, and trace minerals.

To be effective, compost must be organically stabilized, which means that it will not actively ferment. The latter could cause odors, exercise a large oxygen demand possibly resulting in anaerobic pockets, and immobilize nutrients used by the biodegrading microbes. Organic stabilization is achieved by destroying the easily biodegradable (putrescible) fraction of the source material, leaving only very slowly degradable fiber. This occurs in anaerobic digestion as well as in aerobic composting. Important side effects of composting include a lowered C/N ratio and pasteurization (inactivation of human pathogens through exposure to high temperatures below the boiling point of water). Both are achieved in thermophilic anaerobic digestion, but instead of incurring the aeration cost of aerobic composting, which can be substantial, anaerobic digestion results in a fuel gas.

Therefore, most of the composting process occurs inside the anaerobic digester. To be marketable as compost, the dewatered residue merely needs to be further screened and cured. The screening would be geared towards removing some of the undesirable nondegradable organic (plastic) and inorganic (glass, grit) material that has been concentrated in the residue by the digestion process. The curing would consist of short-term aeration to remove odors and effect some further drying. For this study, it was assumed that one-quarter by mass of the filtercake would be rejected in the screening process and landfilled at the prevailing landfill tipping fee. Three-quarters of the wet cake mass is marketed as compost.

2.9 PROCESS ENERGY MODULE

All of the thermal and electric requirements of the total process are calculated in this module, as well as the percentage of gross energy production (in the form of biogas) for each item considered. Reactor temperature, average and design ambient temperatures and thickness of sprayed-on polyurethane insulation are inputs. Based on average and design weather conditions, a detailed thermal balance for the conversion process is calculated. The enthalpies of materials flowing in and out of the biogasification system are calculated for the feed, dilution water, wet biogas, filtercake, and excess filtrate. Conduction, recycled filtrate, and evaporative heat losses are accounted for; conductive losses include those through the walls, roof and bottom of all tanks, as well as piping losses. Radiative heat loss was found to be negligible.

It is important to keep in mind that with a dry feed such as MSW, process heating needs are not a concern. In a properly designed system, these process heating needs can be kept down to around 5% of the energy contained in the gross biogas production, irrespective of digester temperature or climate. Mixing heat input for the CSTR option and metabolic heat production are calculated in this model. They are significant sources of heat, making the biogasification process a net producer of heat in the basecase, where digesters are large and ambient temperatures are temperate.

The biogasification thermal balance calculations are used in the overall energy balance for the facility. The relative quantities of energy (thermal and electric) used by each process module are calculated based on the energy input of MSW as 100 percent. Energy which leaves the system is in the form of SNG (or other energy products), filtercake, and process waste heat. The fuel needed to generate the electricity for the facility is accounted for when power is not generated on site. Overall facility gross and net efficiencies are provided by the results of this module.

Electricity requirements are determined for all of the process modules including preprocessing power, reactor mixing (for CSTR only), pumping in the conversion system, mechanical drying, fans and pumps for the boiler plant, delumper and trommel for composting, energy required by the biogas processing systems, and building utilities. Mixing energy calculations have been revised since the previous model simulation to reflect the results of tests by mixing equipment manufacturers. The new mixing energy values are a factor of 3.5 higher than the previous values at 8% TS. Total heating requirements (or production) in Btu/hr, total electricity in kwh/day, and total percentage of gross biogas production to meet plant energy needs are the results of this module.

2.10 COST MODULE

The Cost Module contains facility cost and credit tables with the costs being broken down by process module. The assumptions section allows input by the user for unit costs such as value of scrap aluminum, compost, and electricity, and cost of land, electricity, boiler fuel, and landfilling. The cost table contains equations to calculate total costs for the components of the process modules. Capital costs are considered long-term or short-term. Annual costs ($/yr) include operation and maintenance (O&M), labor, and fuel/power. Component costs for the conversion and gas cleanup processes were taken from the "Equipment Costs Handbook for Biomass and Waste Systems" compiled by RS&H. Costs for the preprocessing module and residue processing options of boiler plant and boiler/turbine plant were developed in previous versions of the model. The credit table calculates the annual revenue streams for byproducts and waste disposal. Byproducts are scrap iron, scrap aluminum, compost, and excess electricity generated in the cogeneration option. If electricity requirements by the entire system are not met by an on-site power plant, the cost of electricity for each process module is calculated in the cost table. Waste disposal credits are taken for MSW as a function of tipping fee.

2.11 LEVELIZED COST MODULE

The Levelized Cost Module was developed according to the equations supplied by Decision Focus, Inc., (Clark, 1982). This module receives as inputs the costs and credits of the process modules from the cost/credit tables. The user inputs financial parameters such as cost escalation rates and project funding sources. The costs and credits of each process module for long-term capital, short-term capital, O&M, fuel, land rent, byproduct credit, and waste disposal credit are then levelized to produce a net total levelized cost of medium Btu gas, SNG, electricity, or methanol, depending on the biogas processing option being used. The resulting levelized costs include working capital and process development allowances, and represent the price to be charged for the product such that the plant will break even.

2.12 LEVELIZED COST THEORY

The cost analysis of the entire system is based on a levelized cost-of-service price methodology. The cost-of-service price is a price per unit of energy product ($/MMBtu) sufficient to generate revenues to meet the following requirements:

1. Amortize debt,
2. Cover operating and maintenance costs and fuel expenditures, and
3. Provide a return on both common and preferred equity.

The levelized cost-of-service price represents a constant dollar per unit price which, if charged for each unit of output over the life of the plant, would yield the same revenue value as would the actual cost-of-service price, discounted to its present value. Thus, the current dollar cost-of-service price is discounted to its present value and levelized over the life of the plant using a constant dollar annuity factor.

The levelized cost for each process is calculated from the total plant investment, variable operating and maintenance cost, fuel expenditures, income taxes, and working capital. The total plant investment is the sum of all construction costs, including site preparation and improvements; plant and process costs; and indirect costs (sales tax on materials, contingency funds, contractor overhead and fees, engineering and design costs, and cost of spare parts). The total plant investment is allocated to the expected plant output using the product of the service factor and net output, resulting in the specific plant investment expressed in dollars per MMBtu per year.

The variable operating and maintenance cost (VOM) has been estimated for the base year and allocated to the expected output. Working capital (WC) has been estimated as a constant fraction of total revenues received each year. Income taxes (where appropriate) have been computed as a function of revenues and are combined at the state and federal tax rate.

The specific plant investment (SPI) is used with the capital charge rate (CCR) to identify the capital charged to each unit of production for long-term and short-term equipment. The unit variable operating and maintenance cost, the unit fuel cost and the unit land rental cost are then converted to their present worth and levelized using the appropriate annuity factor. The levelized unit working capital factor, adjusted by the weighted-average after tax cost of capital and the tax rate, is determined from the unit capital investment, the levelized unit variable operating and maintenance cost, and the levelized unit fuel expenditures. Credits for byproducts, such as scrap aluminum, are accounted for and may be included in the analysis if desired. A process development allowance (PDA) may also be used; the PDA accounts for an increase in the system costs as they move from one level of development to another, and as the definition of the process becomes more detailed. PDAs are not used for commercial-scale facilities which are similar in size and configuration to previously built facilities, i.e., using established technology. In this analysis, no PDAs are used.

The level unit cost is calculated as the product of the capital charge rate and the specific plant investment, plus the levelized unit variable factors, operating and maintenance, fuel expenditures, land costs, and working capital. A detailed explanation of equations used in the levelized cost model is provided in the 1988 Annual Report.

2.13 OUTPUT SECTION

This section summarizes conversion parameters in addition to net energy production figures. The levelized cost and percentage of total cost for each process module is shown. The levelized credits for MSW and byproducts are also indicated, resulting in a net levelized cost of medium Btu gas, SNG, electricity, or methanol.

2.14 MASS AND ENERGY BALANCE RESULTS

The following table shows the key results of the model for the base case, as well as with various process options selected. Note under "Gas

Utilization" that the system power generating efficiency is well below 30 percent after process power demand has been taken into account.

Table 2: Mass and Energy Products Balance, RefCoM Basecase and Options

1. <u>Base Case</u>	<u>Flow Rate (tons per day)</u>
MSW	500.0
Ferrous metals	20.1
Aluminum	2.3
Plastics enriched stream	21.3
Rejects landfilled	64.5
Water addition	3.5
Recycled filtrate addition (from conversion)	47.5
Refuse derived feed	442.8
Converted to gas (biogas and moisture)	188.5
Filtercake at 50% TS	207.6
2. <u>Options</u>	
Filtercake screening to compost	
Tons compost per day	155.7
Tons screening rejects (landfilled) per day	51.9
Filtercake burning	
Tons quenched ash (70% TS) per day	61.6
Gas Utilization	
4,362,000 scf medium Btu gas (biogas) per day = 2,399 MMBtu medium Btu gas per day	
2,273,000 scf SNG per day = 2,159 MMBtu SNG per day	
111,700 kWh of gas turbine-generated electricity per day = 381 MMBtu of gas turbine-generated electricity per day	
20,900 gallons methanol per day = 1,254 MMBtu methanol per day	

3.0 SENSITIVITY ANALYSES

The sensitivity analysis was performed in the following steps:

1. Establishment of base case.
2. Sensitivity of net gas cost to changes in a single parameter at a time.
3. Sensitivity of net gas cost to digester technology options of SOLCON and SERI High Solids.
4. Sensitivity of energy efficiency to process options and digester technology.
5. Optimization of process options to produce four "best" cases for each digester technology.

3.1 BASE CASE

The base case is defined as a realistic RefCoM system with an optimized retention time and solids concentration. A detailed summary of basecase assumptions was shown in Table 1 in Section 2.0. Process development allowances are not included since this is not assumed to be the first facility built. The parameters which will be varied in the sensitivity analyses are as follows:

Table 3: Parameters Varied in Sensitivity Analyses and Their Base Values

Parameter	Base Case Input
Facility Size	500 tpd
Plant Availability (Service Factor)	0.85
Preprocessing	
Disposition of Plastics	Landfill
Biogasification	
Reaction Rate Constant	0.155 day^{-1}
Gas Utilization	SNG
Post-Processing of Residues	Landfill
Income/Cost Streams:	
Electricity Cost	\$0.05/kwh
Facility Tipping Fee*	\$40/ton
Value of Scrap Aluminum	\$800/ton

The required price of gas for the base case is \$5.0/MMBtu and the base case capital cost = \$40.3 million. The base values for reactor solids content and retention time (SRT) were determined as follows. A reactor solids content of 6.5% TS was artificially selected, then the SRT was reviewed to determine at which SRT the cost is minimal. This optimal SRT was found to be 23 days as shown in Figures 3 and 4. Using this optimum SRT, the TS concentration was varied from three to 12 percent as shown in Figure 5. The net gas cost curve levels out at approximately eight percent TS. This is the highest solids concentration at which RefCoM-type tests have been run, so this figure was used for the basecase, rather than the theoretical optimum of 12% TS.

*Note: Although there is no direct relationship between biogasification charges and landfill charges, it is assumed here that the tipping fee charged by the biogasification facility is equal to the per-ton cost of hauling and landfilling MSW.

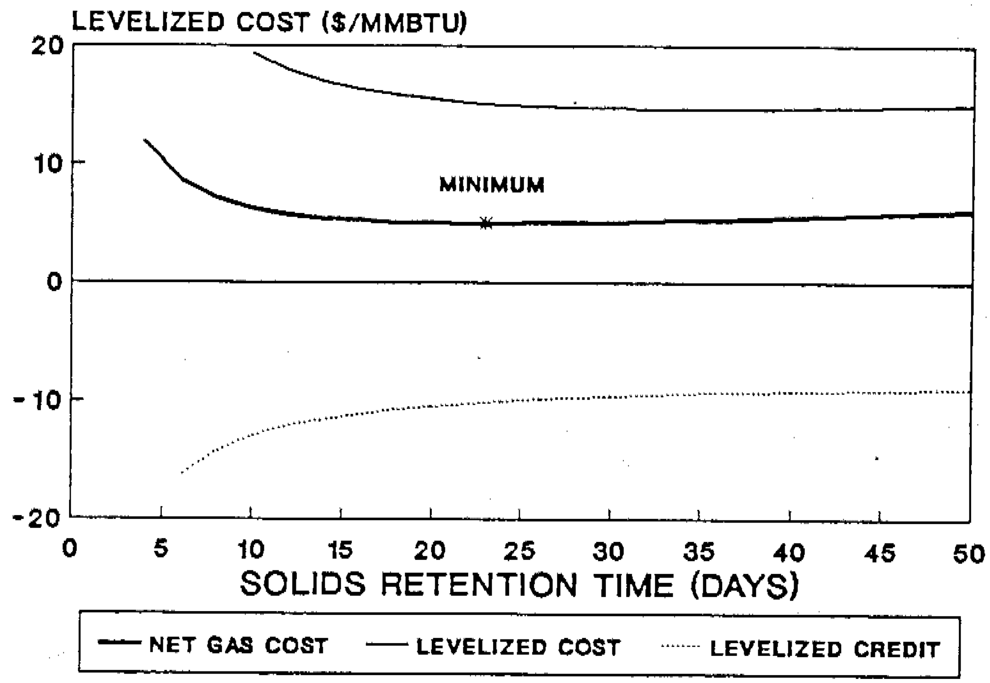


Figure 3: Levelized cost versus solids retention time for the RefCoM base case. The top curve is the total cost to run the facility and service debt; the bottom curve is the income derived from tipping fees and the sale of recyclables and fuel gas (income expressed as a negative cost). When these revenues are subtracted from the total costs, the middle curve is generated; it represents the net price of gas that must be charged to meet economic objectives. These costs are expressed per unit of gas; at short retention times gas production is low and per-unit costs are high, which is why the curves veer off sharply below ten days retention time.

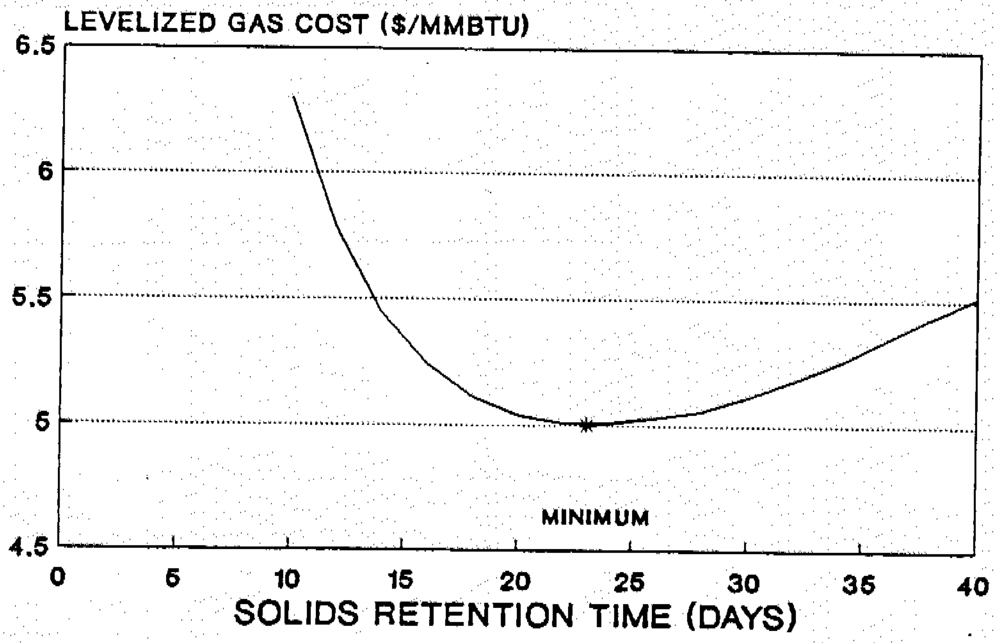


Figure 4: Magnified section of the net gas cost (middle curve) of Figure 3, to illustrate optimum retention time. The lowest gas cost, $5.0/MMBtu is achieved at 23 days SRT for the base case. Any retention time between 14 days and 40 days will yield a gas cost within ten percent of the optimum.

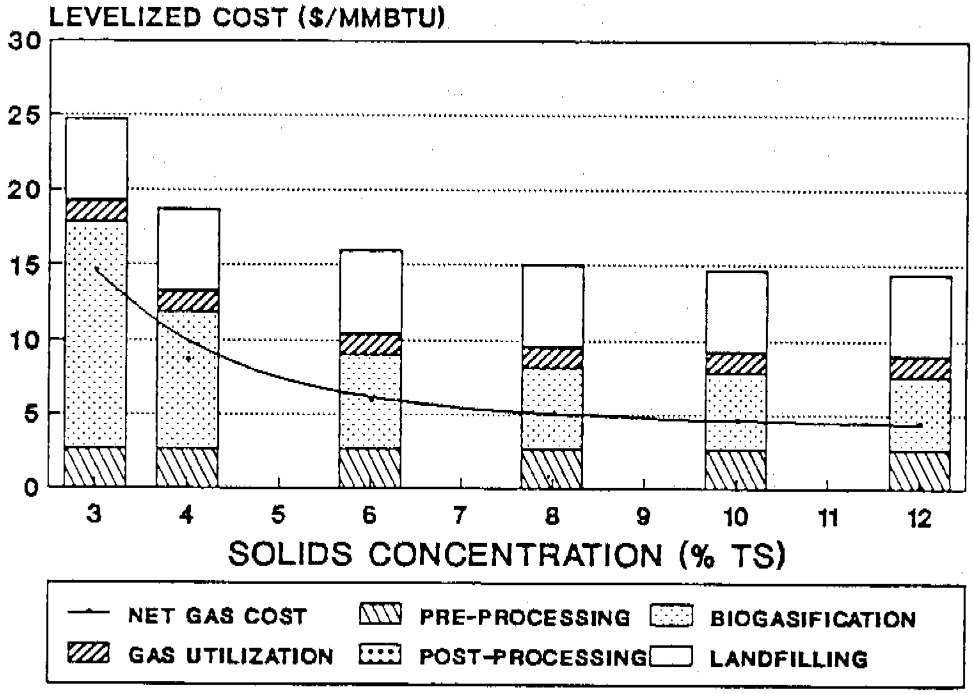


Figure 5: Sensitivity of levelized gas cost to solids concentration (TS%) for the RefCoM base case. The costs (operating and debt service) attributed to each facility module are stacked up; revenues from tipping fees and sale of recyclables are subtracted, resulting in the net gas cost curve. This curve on the graph is the locus of all the gas prices. At 8% TS for example, the price is $5/MMBtu. Very dilute reactors require large volumes, hence their high cost. At higher concentrations, the increased mixing cost negates the gains from reduced volumes. Note that there is no proof that RefCoM can be operated above 8% TS, because of bridging effects in the slurry that prevent complete mixing.

Therefore, base case SRT is 23 days and TS concentration is eight percent. These values were then used in generating Figures 3, 4 and 5.

Note the bar/line chart format in Figure 5. This format is used throughout the sensitivity analyses. The bars show levelized cost by process module, and the curve shows net levelized gas cost (net levelized gas cost = levelized costs - levelized credits). The credits (tipping fee and sale of recyclables) are equal to the distance between the tops of the stacked bars and the curve. Figure 6 shows a detailed view of the levelized cost distribution by process module and by cost type. Debt service is payment on capital cost and financing. Recurring costs include O&M, fuel and power, land rent, and working capital costs. Figure 7 shows the levelized incomes which are required to offset the costs. The tipping fee accounts for 60 percent of the income, with SNG being sold at $5/MMBtu. Capital costs are shown in Figure 8; the proportions are quite similar to the levelized costs (which include recurring costs) as shown in Figure 6. The total projected capital cost is $40.3 million.

3.2 SINGLE PARAMETER COST SENSITIVITY

The cost sensitivity analyses were performed by varying a single parameter over a range of values. These parameters are grouped according to process module: preprocessing, biogasification, gas utilization and postprocessing of residues.

3.2.1 General

Facility size and availability are parameters which apply to the entire system rather than specific modules. Figures 9 and 10 show the effects of varying the size of the facility on the levelized gas cost. Figure 9 shows costs by process module and Figure 10 shows costs by type--debt service or

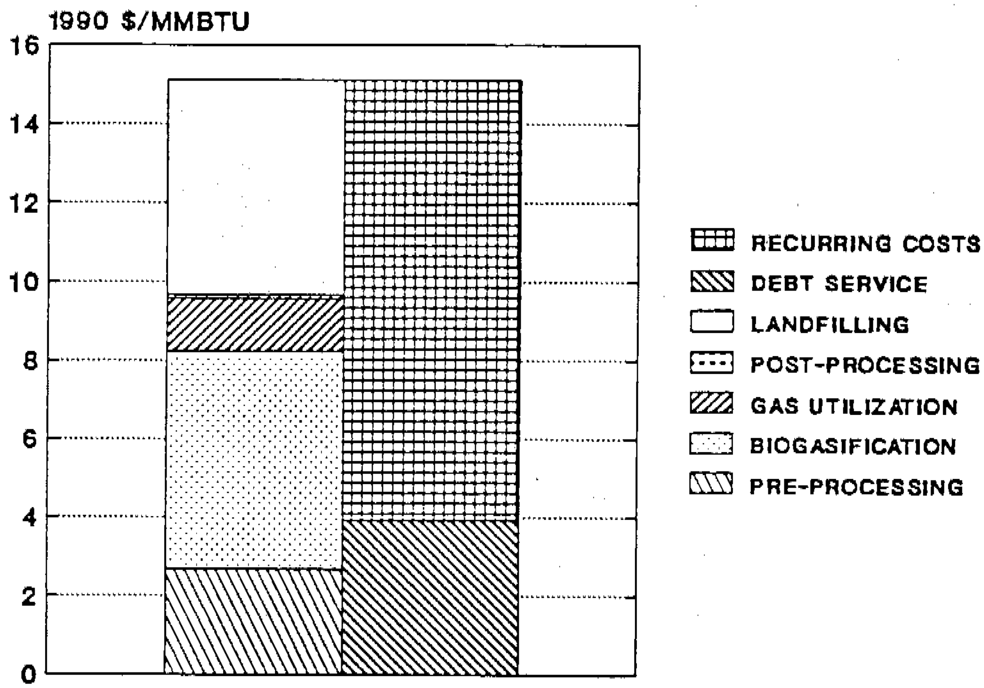


Figure 6: Cost distribution for the RefCoM base case. In the left half, the cost contribution of each module is added up; in the right half, recurring costs (operation, maintenance, power, fuel, land rent, and working capital) are compared to the cost of debt service.

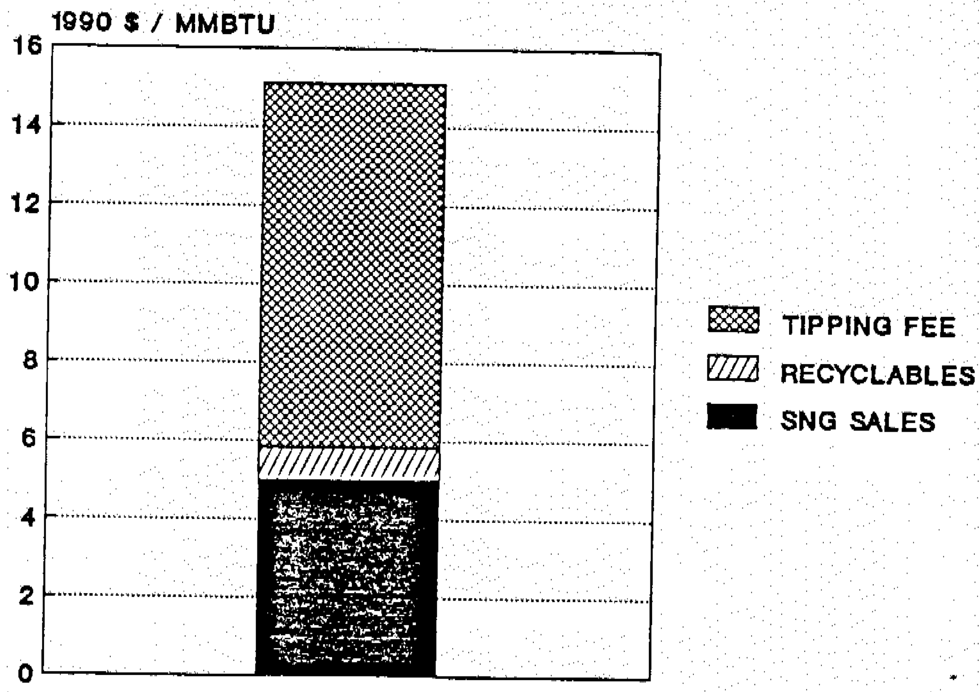


Figure 7: Revenue distribution for the RefCoM base case. The tipping fee ($40/ton MSW) accounts for approximately 60 percent of the income, with SNG being sold at $5.0/MMBtu.

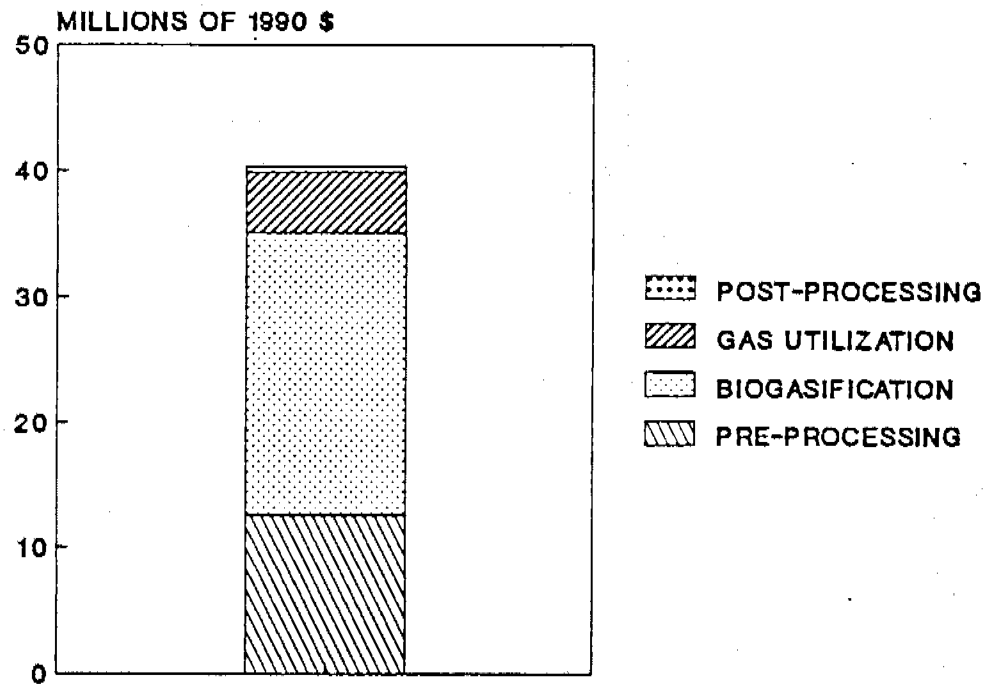


Figure 8: Capital cost by module for a 500-tpd RefCoM plant. The proportions are similar to the levelized costs, which include recurring costs.

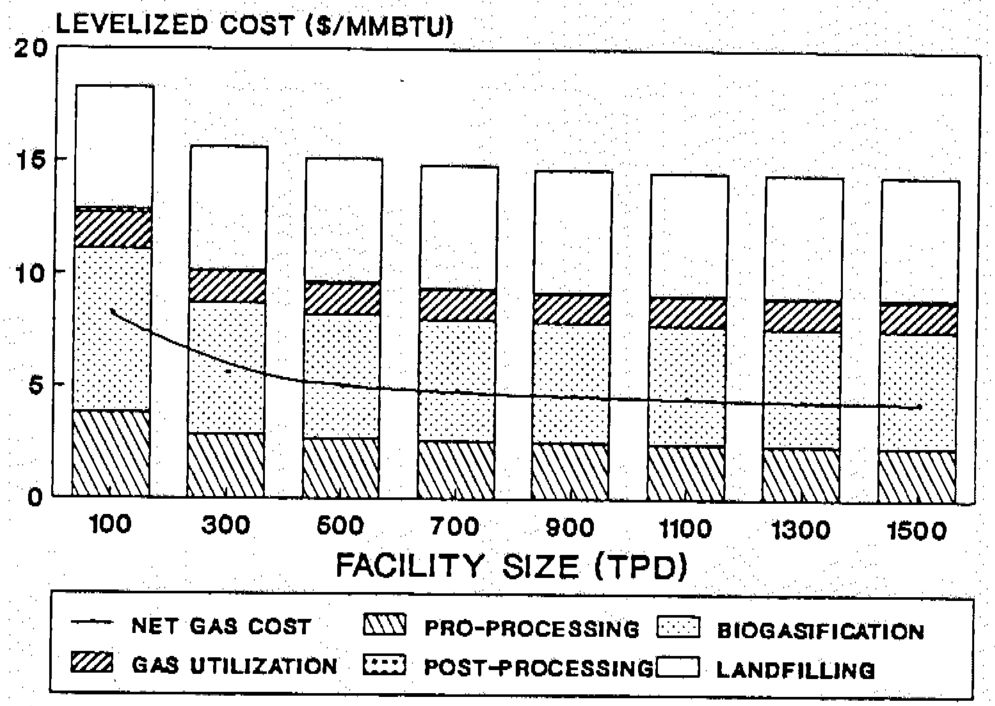


Figure 9: Levelized cost of gas ($/MMBtu) versus facility size (tpd MSW capacity). Base case = 500 tpd. The total cost associated with each process module is shown in the stacked bars. Revenues from tipping fees and the sale of recyclables are subtracted to yield the price at which gas must be sold to break even (curve).

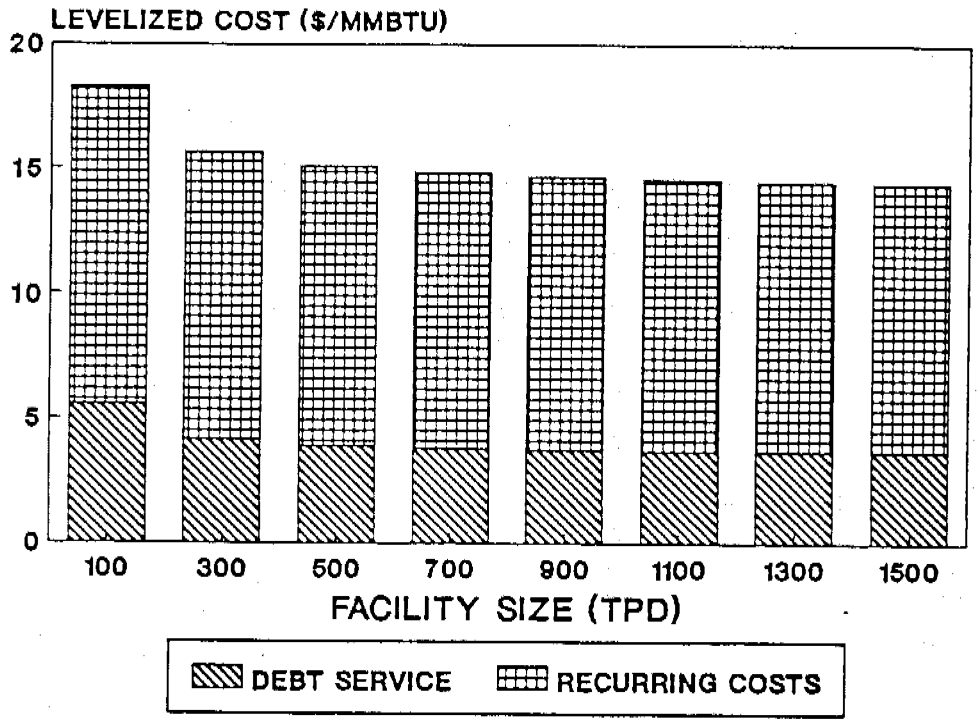


Figure 10: Levelized gas cost versus facility size. Similar to Figure 9, only here the proportion of debt service to all other operating costs is illustrated.

recurring costs. Note on Figure 9 that the cost of landfilling residue is constant and independent of scale because the same conversion efficiency is assumed throughout, therefore the same proportion of MSW input ends up as residue to be landfilled. Economies of scale are appreciable up to 500 t d^{-1} and are found mainly in the biogasification and preprocessing modules, at least for the base case, where all residue is assumed landfilled. Should an expensive postprocessing technology such as combustion be selected, stronger economies of scale should appear. From Figure 10, it can be seen that operating costs consume most of the budget.

The effect of facility availability on levelized costs is shown in Figure 11. The service factor or availability is the fraction of the time that the facility is actually running. A service factor of 0.9, for example, means that the plant experiences ten percent scheduled and unscheduled downtime or five weeks per year. If the facility is not operating for substantial periods of time, gas production will be reduced, which increases levelized costs since the latter are calculated by dividing total costs by gas production. Note that no penalty is taken into account for disposing of garbage during downtime. Note also that the landfilling costs are independent of the service factor, as in Figure 9. The net gas cost continues to decrease as the facility is operated more reliably.

3.2.2 Preprocessing

The disposition of the plastics-enriched stream (see Section 2.3) is evaluated as a preprocessing parameter. In Figure 12, the effect of plastics disposal options on the levelized gas cost is illustrated for four residue disposal options. Plastics may be landfilled, recycled (no net revenue) or burned. As one can see from the figure, burning plastics is not considered a viable alternative when solid residues are landfilled or composted; a separate

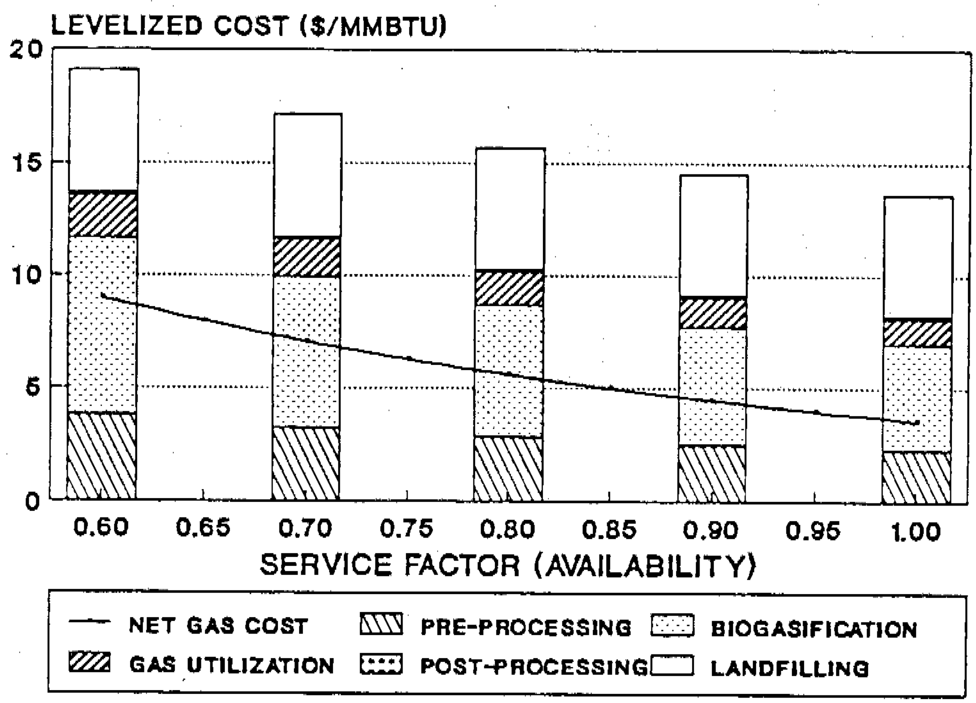


Figure 11: Levelized gas cost versus service factor. Base case service factor = 0.85. The service factor is the fraction of the time that the plant is actually running. Cost display as in Figure 5.

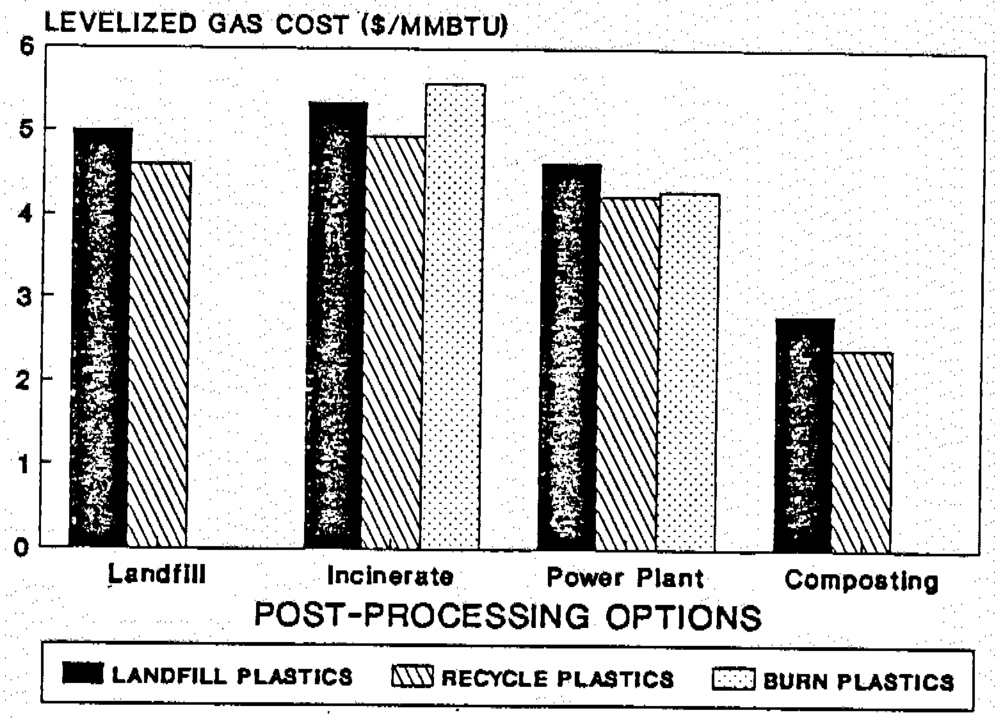


Figure 12: Levelized gas cost versus disposal strategy for the plastics-enriched stream. It can be landfilled, burned, or recycled (at no net cost or income); base case = landfilled. Four disposal strategies are considered for the final solid residue left over after anaerobic digestion.

boiler is not included for burning plastics only. The recycling option is more economical than landfilling or burning for all postprocessing alternatives--landfilling, incineration, power generation and composting of solid residues. This is so because it was assumed here that recycling constitutes disposal at no net cost, whereas costs are incurred for burning or landfilling.

3.2.3 Biogasification

A sensitivity parameter which pertains to the biogasification process is the reaction rate. Figure 13 shows the effects of varying the reaction rate constant on the levelized gas cost. An increase of the rate constant from 0.10 to 0.25 day^{-1} produces significant savings; the gas cost is nearly cut in half from \$7/MMBtu to \$3.60/MMBtu. Savings are moderate for reaction rates greater than 0.25 day^{-1}, which are highly speculative anyway. The bar graph portion of the figure shows that reductions occur mainly in biogasification and landfilling. Biogasification is affected since more efficient conversion results in smaller reactor volumes and consequently lower energy expenditures for mixing. Since the retention time is kept constant at 23 days, increased reaction rates are accompanied by increased conversion. More extensive conversion to gas means less residue remains to be landfilled. Finally, since the gas production increases, the cost per unit of gas is also reduced. Note that this analysis is conservative and the cost reduction will be more pronounced if for every rate constant the optimal SRT is used (23 days is only optimal for $k = 0.155\ day^{-1}$).

3.2.4 Methane Enrichment Digestion

Production of high-Btu gas using the MED process, as described in Section 2.5.4, is evaluated here as a sensitivity parameter related to gas utilization. The effects on levelized costs can be seen in Figure 14 where

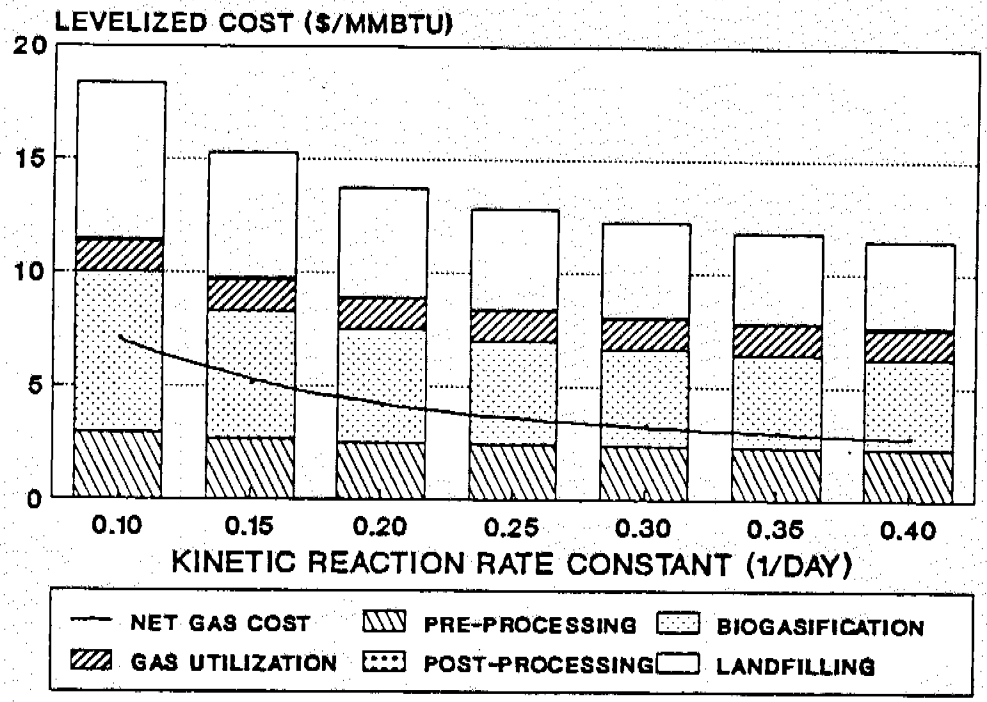


Figure 13: Levelized gas cost versus first order reaction rate constant. Base case = 0.155 day^{-1}. Cost display as in Figure 5: costs per module are stacked up, revenues from tipping fees and recycling are subtracted to yield required gas price (curve).

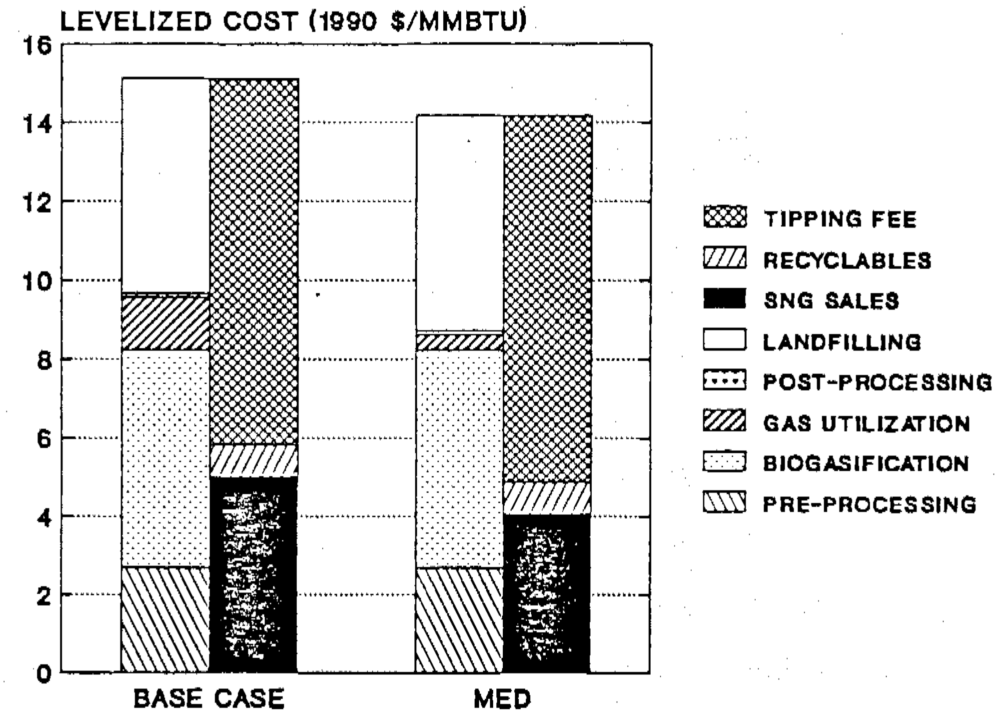


Figure 14: Levelized gas cost versus SNG production methods. The base case shows costs and credits by module for gas cleanup using a membrane separation system. The costs for purification by MED are nearly $1/MMBtu lower than the base case, which is reflected in a lower SNG price to break even.

gas cleanup using membranes (SNG option) and using MED are compared. The net gas cost including MED cleanup is estimated to be \$4.05/MMBtu, whereas the gas cost with membrane purification is projected to be \$5.00/MMBtu in the base case. It must be emphasized that this is only a first attempt at estimating the cost of MED, based on the assumptions listed, and that a more thorough evaluation is necessary.

3.2.5 Gas Utilization and Postprocessing of Residue

Various combinations of gas utilization and residue post-processing options are shown in Figures 15 through 18. In each figure a particular gas utilization option is chosen and four solid residue processing options are contrasted: direct landfilling, incineration, combustion with power generation, and screening to produce compost (one-quarter by mass was rejected and landfilled). As explained in Section 2.0 (see also Figure 1), four gas utilization options are considered: marketing of medium Btu gas, marketing of SNG (pipeline quality gas), conversion of medium Btu gas to electricity using a gas turbine and sale of this power, and conversion of the biogas to methanol.

The following general comments can be made about Figures 15 through 18. For all gas utilization options, the levelized energy product cost is the lowest if residue is refined to compost. Gas costs for landfilling, incineration and power plant options are substantially higher. The total cost of the power plant option is approximately the same as or less than that for composting of residue, with landfilling and incineration of residues having substantially higher costs. The power plant option is dominated by post-processing (= burning) costs, but biogasification costs are greatly reduced since power is generated on site to satisfy the large energy demand for mixing. It is important to remember that this analysis is based on the RefCoM process, which is relatively energy-inefficient. Should a more energy efficient process be substituted, all energy breakeven costs would come down substantially.

When medium Btu gas is the final product (Figure 15), gas utilization costs are low since only minimal gas processing occurs. Power plant and composting costs are about the same, with higher post-processing (burning) costs for the power plant being offset by higher digestion costs for composting. By comparing composting with landfilling, one can see that the additional cost for the composting process (post-processing) is a worthwhile investment since it reduces landfill costs dramatically, resulting in a much lower net levelized gas cost for composting. Incineration is not a viable alternative economically since post-processing costs are significant compared to the reduction in material to be landfilled.

Costs for SNG production are shown in Figure 16. Trends are similar to those described for medium Btu gas options, except that gas utilization (= gas processing) costs are higher. Power plant costs are lower than composting costs since the energy costs for gas cleanup are significant when power is not generated (gas cleanup to pipeline quality consumes a lot of compression power). Here again, composting proves superior to landfilling of residues both in total levelized cost and net gas costs. The additional cost for incinerating residues is not yet justified at this tipping fee ($40/ton), especially since the biogasification facility is thermally self-sufficient and does not need any additional heat from the incineration process.

Figure 17 shows post-processing of residue options when biogas is used to produce electricity in a gas turbine. Here again, trends are similar between the post-processing options: incineration and landfill costs are the highest, followed by composting and power plant. Gas utilization is a larger

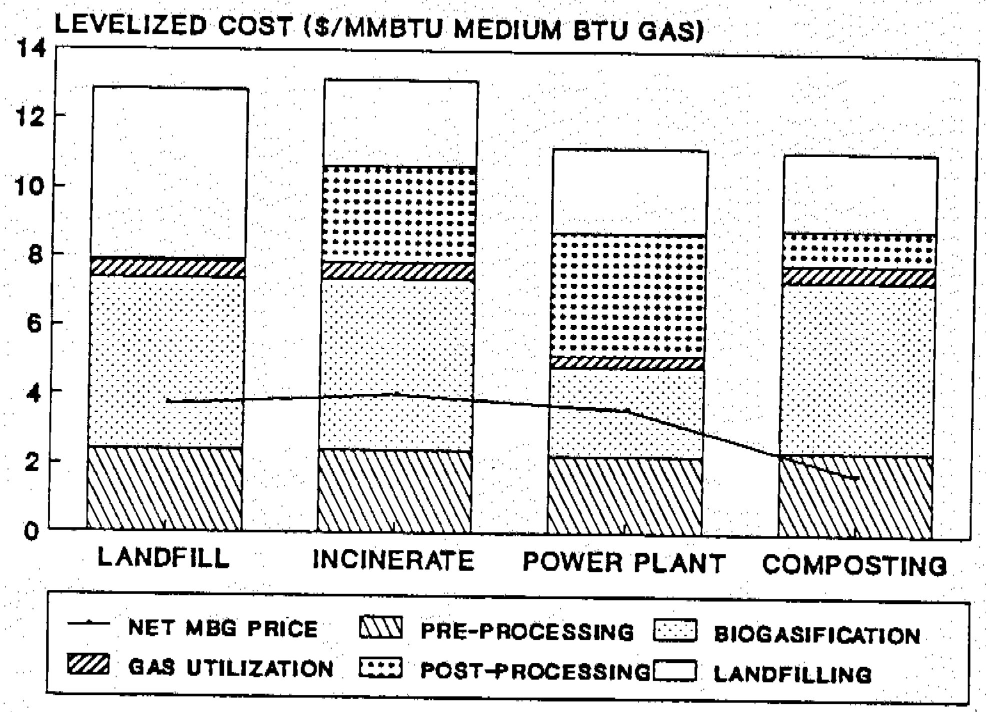


Figure 15: Levelized medium Btu gas cost versus solid residue disposal options. Four disposal options are considered: landfilling all of it, incineration, combustion with power generation, screening to produce compost. The curve indicates the gas cost necessary to break even.

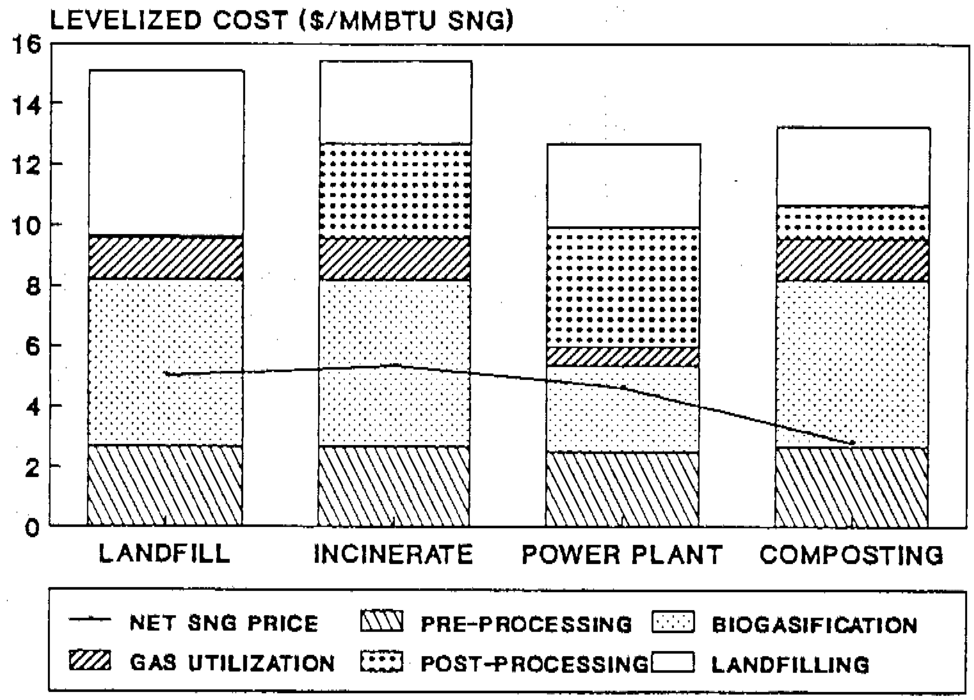


Figure 16: Levelized cost of SNG (substitute natural gas) versus solid residue disposal options. Base case = landfilling all solid residue. The curve indicates the gas cost necessary to break even.

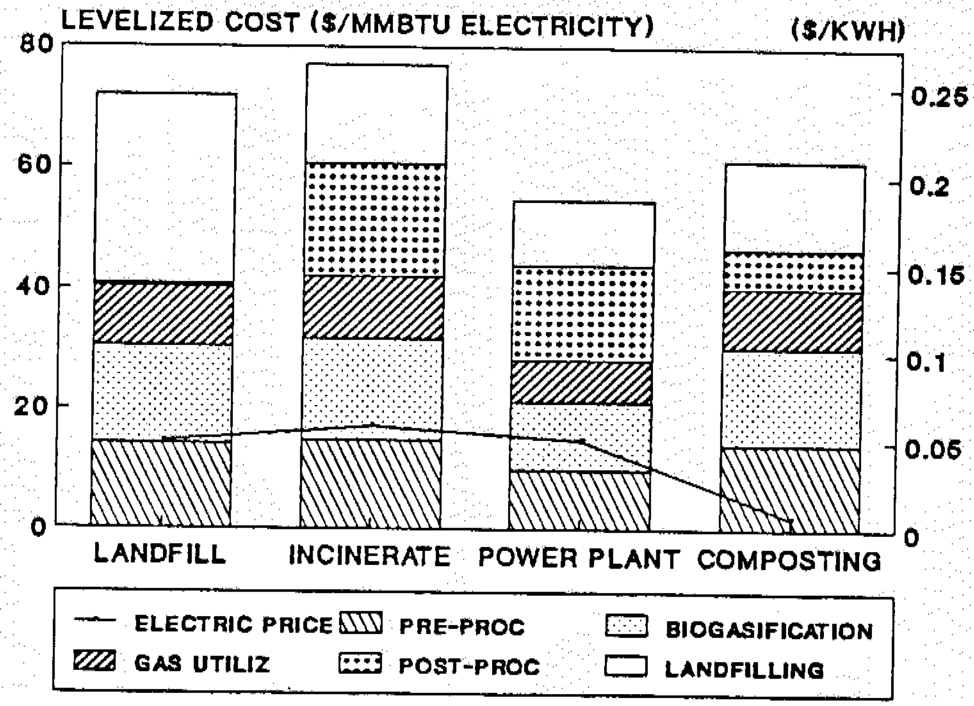


Figure 17: Levelized cost of electricity generated with a biogas-fueled turbine versus solid residue disposal options. The curve indicates the price of electricity required to break even.

portion of the total costs due to the expense of the gas turbine and associated equipment. The electricity must be sold at approximately $0.05/kwh for landfill, incinerate, and power plant options but if residue is composted, electricity sale at $0.01/kwh is sufficient for the facility to break even.

When methanol is the energy product, as shown in Figure 18, gas utilization costs make up an even larger percentage of the total costs than for medium Btu gas, SNG, or electricity. The total costs of the facility with landfilling or incineration of residue are about the same; costs for power plant and composting are about the same also, but they are lower. The methanol must be sold at approximately $0.70/gal for all options except for composting, which only requires a $0.50/gal selling price.

3.2.6 Income/Cost Streams

The sensitivity of levelized gas cost to changes in electricity cost, tipping fee, and value of recyclables is shown in the next figures. As the cost of electricity increases, the net levelized gas cost increases linearly, as shown in Figure 19. The greatest impact of increased electric rates is seen in the biogasification module because it requires large amounts of power for mixing. Gas utilization costs increase also, due to the energy required for compression in gas cleanup to SNG.

Figure 20 shows the effect of increased tipping fees on the levelized gas cost; it is a linear relationship. The top line shows facility costs increasing, but the bottom line shows credits increasing more rapidly (credits are shown as negative costs). This results in a net gas cost (middle line) which decreases steadily as the tipping fee increases, such that tipping fees of $90/ton require no revenue from gas sales.

The value of recyclables can have a significant impact on the net gas cost, as shown in Figures 21 and 22. Eight recycling cases were evaluated, as

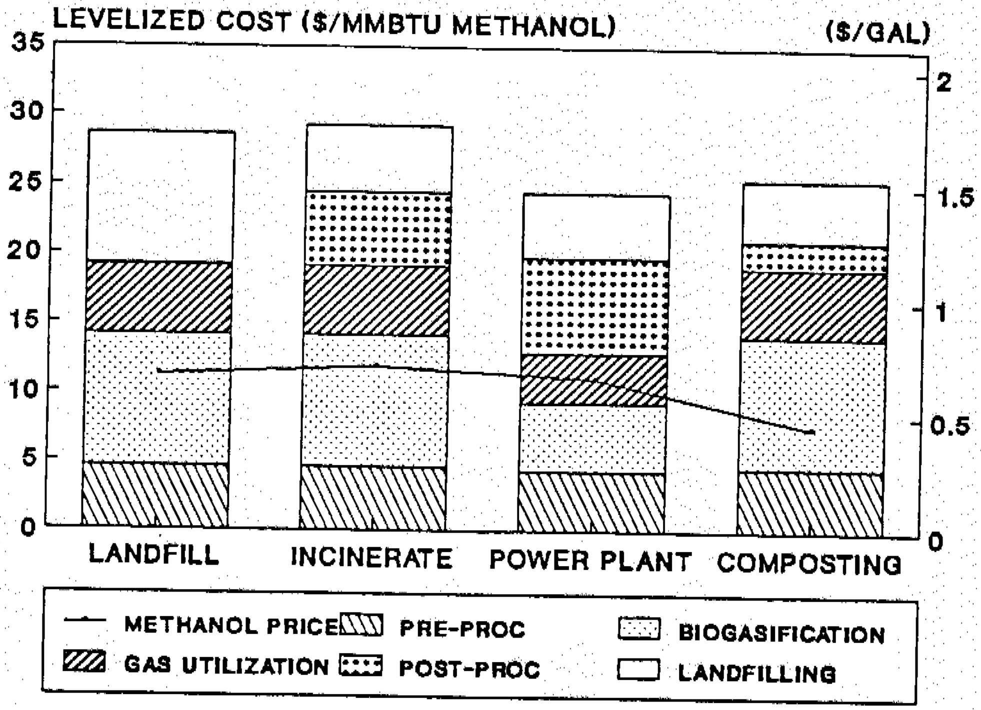


Figure 18: Levelized cost of methanol produced from biogas versus solid residue disposal options. The curve indicates the price of methanol required to break even.

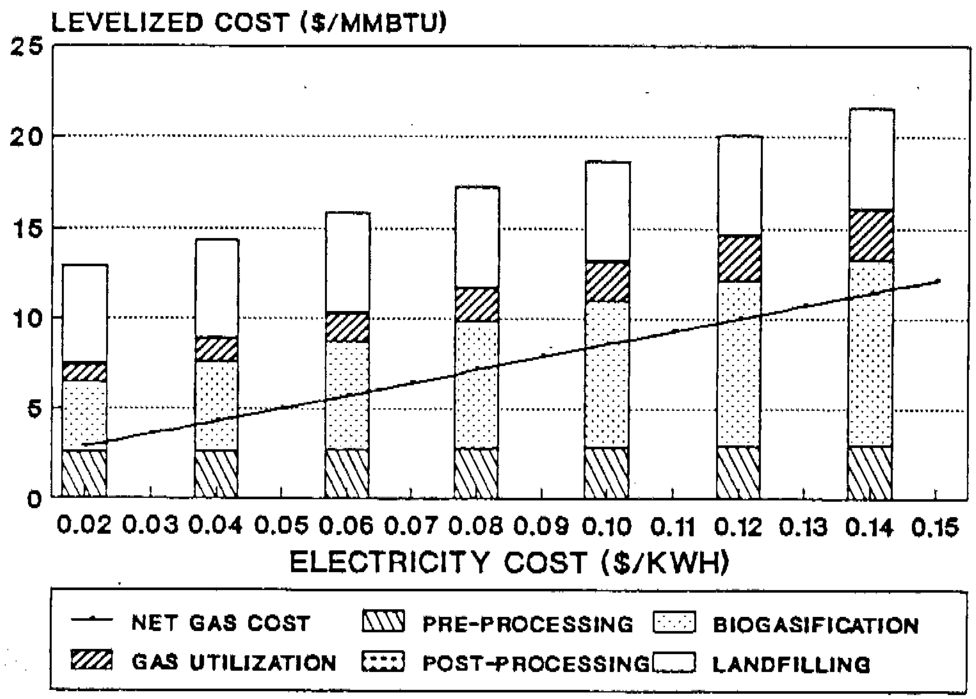


Figure 19: Levelized gas cost versus electricity purchase rate. Base case = 5¢/kwh. RefCoM is a power-hungry process so costs are sensitive to the price of electricity.

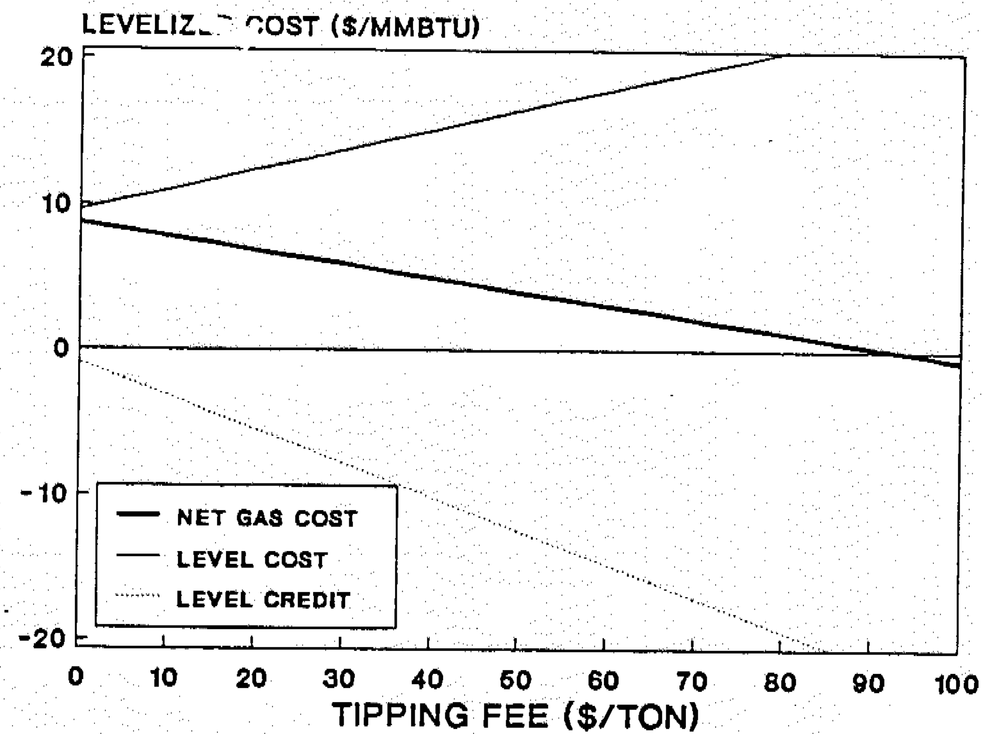


Figure 20: Levelized gas cost versus tipping fee. Base case = $40/ton. This is both the tipping fee charged by the anaerobic digestion facility and the tipping fee charged by the landfill. The top line is the total levelized cost, the bottom line is total credits (expressed as negative costs). Subtracting credits from costs yields the middle line, namely net gas cost.

defined in Table 4. From Case 1 to Case 8, increasingly favorable economic conditions are considered.

Table 4. Definition of Recycling Cases

Case No.	Disposition of Plastics	Disposition of Residue	Value of Recyclables in $/Ton Ferrous	Value of Recyclables in $/Ton Aluminum	Value of Recyclables in $/Ton Compost
1	landfill	landfill	-40	-40	--
2*	landfill	landfill	0	800	--
3	landfill	compost	0	800	-10
4	landfill	compost	0	800	5
5	recycle	compost	0	800	5
6	recycle	compost	100	800	5
7	recycle	compost	100	2,000	5
8	recycle	compost	100	2,000	20

*Base case

In the first case, ferrous metals and aluminum are landfilled; negative values are costs, i.e., -$40/ton is the landfill tipping fee. In case 3, the residue is composted and given away, but $10/ton transportation costs are incurred. Remember that it is assumed that one-quarter by mass of the dewatered digested residue is rejected in the screening process and landfilled; three-quarters by mass survives the screening as marketable compost. The values of recyclables listed bracket actual market values in the U.S. in 1989. Some prior removal of aluminum cans from the waste stream is assumed, lowering the MSW aluminum content to 0.45% by mass. Plastics recycling is at zero cost or income.

Figure 21 shows the credits for each of the recyclable products, as well as the net levelized gas cost for each of the cases described above. Note that when residue is composted and sold at $20/ton, and ferrous metal and aluminum can be marketed at high prices (case 8), it is not necessary to sell

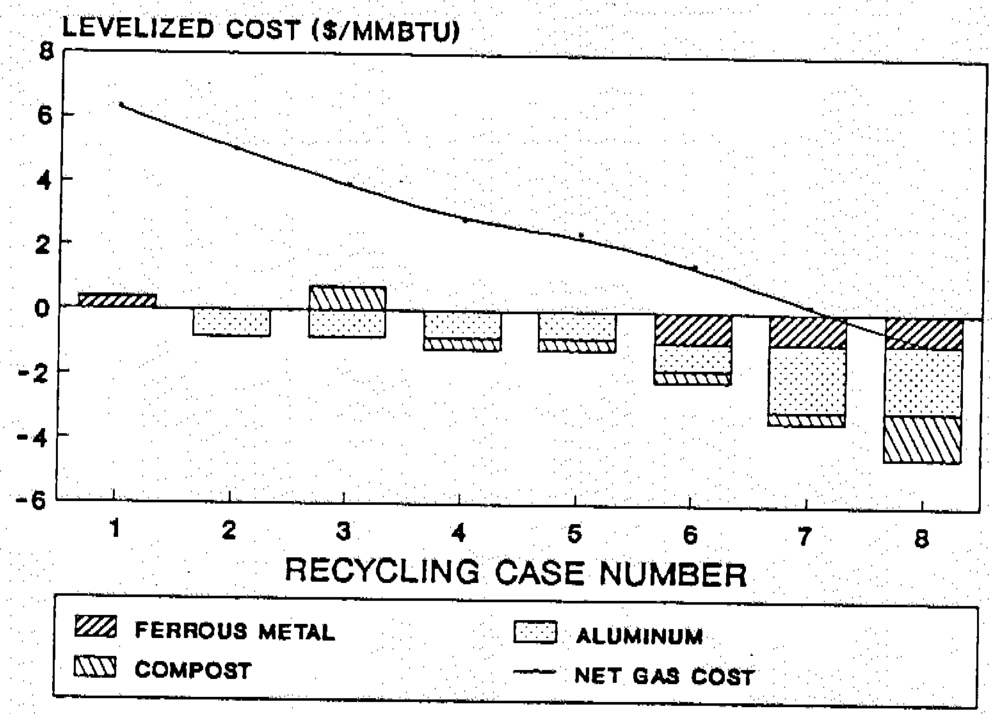


Figure 21: Levelized gas cost versus eight recycling scenarios (see text). Base case = Case 2. The curve indicates the levelized gas cost for each case considered. The bar diagrams indicate costs (positive) and income (negative) resulting from recycling.

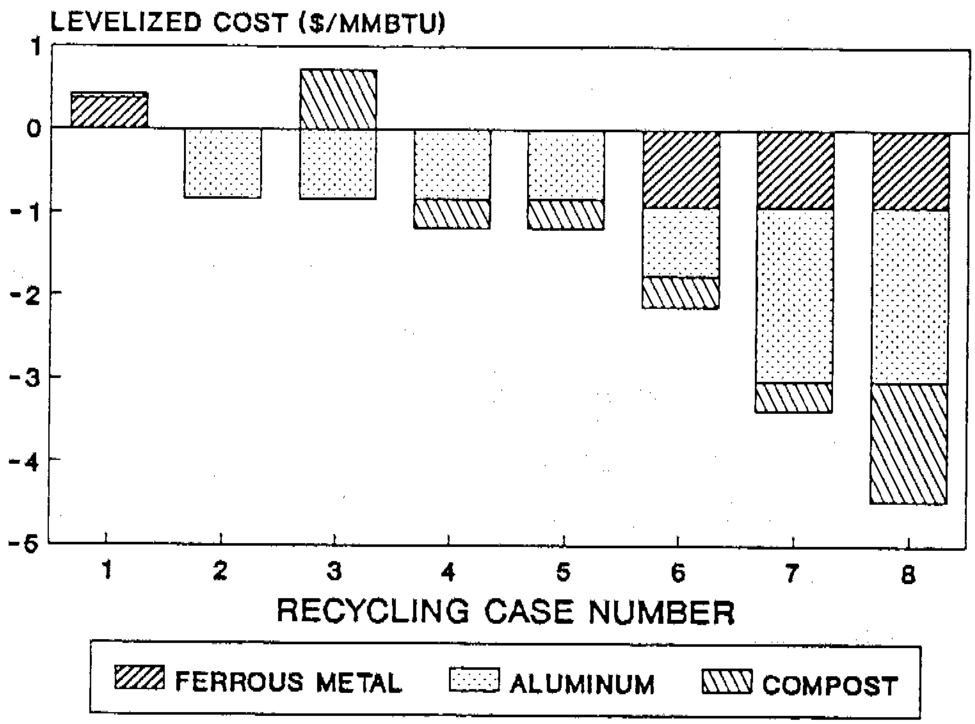


Figure 22: Costs and credits related to recycling for eight scenarios (see text). Base case = Case 2. This is merely a magnified section of Figure 21.

the SNG; the levelized gas cost is negative. Figure 22 shows the detail of the credits (and costs).

3.3 DIGESTER TECHNOLOGY COST SENSITIVITY

3.3.1 General Comments

Three digester technologies are considered here:

- Low solids continuously fed and mixed, prototype: RefCoM ("Refuse Conversion to Methane");
- Low solids continuously fed unmixed, prototype: SOLCON ("Solids Concentrating digester");
- High solids continuously fed and mixed, prototype: SERI HS ("SERI High Solids digester").

RefCoM is an impeller-mixed CSTR demonstrated at a scale of up to 21 tpd RDF in Pompano Beach, Florida. SOLCON is a process developed by IGT in which suspended solids separate in a float layer and liquid is withdrawn from the liquid layer underneath. This stratification causes the solids to remain in the reactor longer than the liquid (SRT larger than HRT). A pilot SOLCON plant is in operation at Walt Disney World. SERI HS is operated in the solid phase, up to 40% TS, and is mixed very slowly with a device developed for such high viscosity material. Three large bench-scale reactors of this type are operated at SERI.

An economic and engineering comparison of these three technologies was made using most of the base case assumptions listed in Tables 1 and 2. The following variables may depart from the base case: reactor TS content, SRT, SRT/HRT ratio, reaction rate, and conversion performance. In all cases, it is assumed that the plastics-enriched stream is landfilled, the biogas is upgraded to SNG, and the solid residue is entirely landfilled.

For a detailed description of RefCoM assumptions and costs, see Sections 2.14 and 3.1. The SOLCON and SERI HS technologies will now be discussed in sections 3.3.2 and 3.3.3, respectively.

3.3.2 SOLCON

An average effluent TS content of four percent was selected, corresponding to 8.7% TS influent which is approximately the highest solids concentration that was successfully demonstrated at the Walt Disney World site. Above this value the solids stratification which is essential for this process breaks down. SOLCON is characterized by an SRT/HRT ratio larger than unity; how this ratio affects economics is illustrated in Figure 23 where cost versus SRT is displayed for four values of the SRT/HRT ratio. A ratio of 1 means the conversion is equal to that of a CSTR; yet no mixing energy needs to be expended. Note from Figure 23 that the higher the SRT/HRT ratio, the longer the optimal SRT (= the SRT that results in the lowest cost). For the SOLCON base case, a ratio of 2.0 was chosen; it may seem high compared to the pilot results, but is justified because of the very buoyant nature of RDF and the larger depth of full-scale systems, which should foster better stratification. At this SRT/HRT ratio, the optimum SRT is 51 days, corresponding to an HRT of 26 days.

3.3.3 SERI HS

Average HRT is found by dividing reactor volume by volumetric influent rate. Average SRT is found by dividing reactor volume by volumetric effluent rate. At high solids content, the reacting material experiences a significant volume reduction during its stay in the reactor. Consequently, the volume of effluent is smaller than the volume of influent and the SRT/HRT ratio is larger than 1. The MSWAD model is not designed to automatically deal with these situations, so the following procedure was followed:

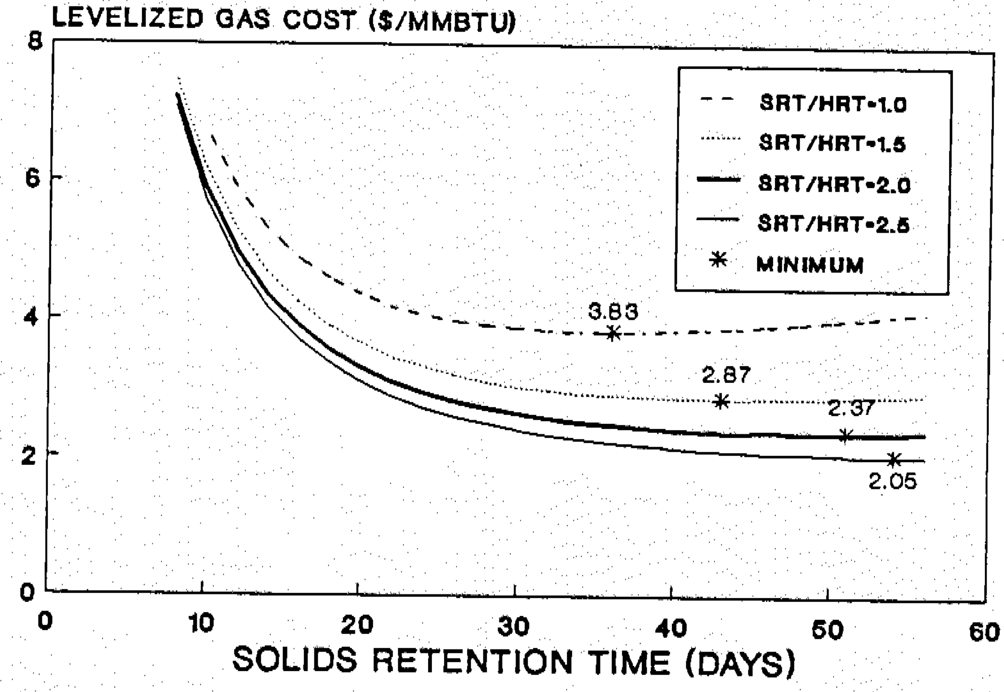


Figure 23: Levelized gas cost ($/MMBtu) versus solids retention time (SRT) in days for four different SRT/HRT ratios, using the SOLCON unmixed digester. The cost-optimal SRT for each SRT/HRT ratio is indicated on the graph (asterisks) together with the corresponding cost.

a) enter an SRT/HRT value larger than 1;

b) increase SRT to keep calculated HRT at the same level;

c) run model;

d) calculate SRT/HRT ratio from effluent tpd/influent tpd;

SRT/HRT = effluent volume/reactor volume ÷ influent volume/reactor volume

= effluent volume/influent volume

= effluent mass/influent mass

e) compare with assumed SRT/HRT;

f) iterate until the SRT/HRT ratios agree.

It is important to have an internally consistent calculation because reactor volume is calculated based on HRT, while conversion performance is based on SRT.

Conversion kinetics were modeled after the actual experiments at SERI. A set of inputs closely paralleling the experiments was used, then the reaction rate constant was adjusted until the outputs were in close agreement with the experimental data (see Table 5). The best match was found with a reaction rate constant of 0.1 day^{-1}, a good rate for a mesophilic (35°C) reactor. A rate of 0.155 day^{-1} was estimated for RefCoM, a process operated at 60°C.

Table 5: Estimation of First Order Reaction Rate Constant for SERI High Solids Reactor

Parameter	Model Input	Experimental Data
TS% inside reactor	35% TS	35% TS
HRT (days)	21	21
bulk dry matter density	16.8 lbs TS/cu ft	---
	270 g TS/L	---
SRT/HRT	1.60	---
SRT (days)	33.5	---
% CH_4 in biogas	65%	66%
	Model Output Assuming k = 0.1 day^{-1}	
VS conversion %	71%	80-90%
Methane yield	4.96 scf CH_4/lb VS_a	6.25
	309 mL CH_4/g VS_a	390
v/v/d biogas	9.3	9.6
v/v/d methane	6.0	6.3
VS loading rate	1.21 lb VS/cu ft/day	0.94-1.25
	19.4 g VS L^{-1} day^{-1}	15-20

The model output and the experimental results diverge for two parameters: VS conversion and methane yield. This is probably due to the fact that the RDF composition listed in Table 1 was used in the model; it is different from the RDF composition measured in the experiments. An additional reason for the discrepancy may be that SERI HS was modeled here as a continuous feed CSTR, while in fact it is intermittently fed, which should lead to better average conversion since short-circuiting is reduced. The TS density in Table 5 was chosen based on RDF densities measured at SERI, as well as some independent results from the University of Florida and a pilot DRANCO facility operated at Walt Disney World. The exact value of this parameter remains elusive; its impact on cost is illustrated on Figure 24. It was observed that costs level off above 28% TS if the dry matter density is set at 16.8 dry lbs/cu. ft. Consequently, there is no benefit in using higher TS contents, which may inhibit the microbiota anyway, and 28% TS was used for the

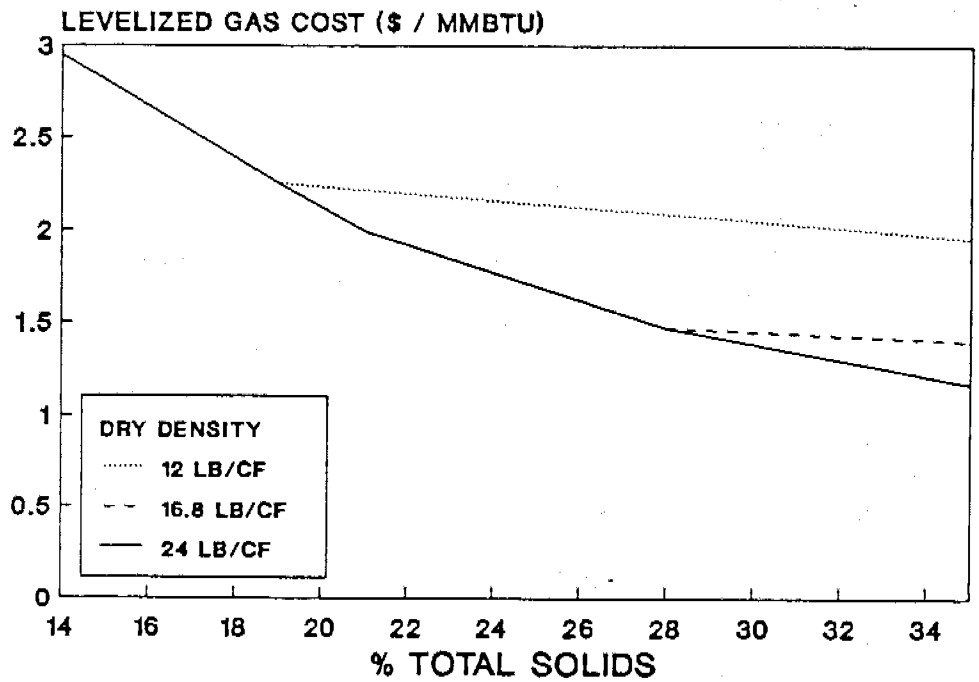


Figure 24: Levelized gas cost versus reactor (=effluent) dry matter concentration for a high solids continuously mixed reactor (CSTR). Three possible dry matter densities (lbs TS/cu ft or grams TS/liter) are considered. Under 19% TS, all three yield the same cost and their curves coincide. Between 19 and 28% TS, the curves for the two highest densities coincide.

SERI HS base case. Under those conditions, the cost-optimal SRT is 67 days with an SRT/HRT ratio of 1.68 resulting in an HRT of 40 days. The key parameters used in comparing the three technologies are listed in Table 6.

3.3.4 Comparison of RefCoM, SOLCON, and SERI HS

Costs and incomes for the three technology base cases as described in the previous sections are displayed on Figure 25. The display is consistent with previous figures in this report: levelized costs expressed in \$/MMBtu are stacked up in the left half of the bar diagrams, whereas the income streams required are added up in the right half. RefCoM is handicapped by its high mixing costs, but the exact mixing requirements with state-of-the-art mixers still have to be confirmed. Gas cost is \$5.0/MMBtu. SOLCON does not suffer from this handicap, but being very dilute, it requires large digester volumes. Required gas cost is \$2.37/MMBtu.

SERI HS has minimal reactor volume and low mixing energy requirements (0.4 hp/1000 gal or 79 W m^{-3} was used), resulting in the lowest gas cost, \$1.48/MMBtu. Note also that the RefCoM costs are all "stretched out," due to the fact that at its economic optimum, less gas is produced (see Table 6) which increases the cost per unit of gas.

Cost versus reactor TS is displayed in Figure 26 for all three technologies. Note that effluent TS% is in the x-axis. RefCoM values above 8% TS are speculative, as are SOLCON values above 4% TS (corresponding to 8.7% TS in the influent).

The three technologies are at different levels of development: RefCoM is the most proven, having been demonstrated at 21 tpd RDF. SOLCON has only been operated at 0.05 tpd (two orders of magnitude smaller), whereas SERI HS is run at 0.001 tpd (four orders of magnitude smaller). It is unclear how solids would be distributed and managed in a full-scale SOLCON reactor. The

Table 6. Parameter Values used for Biogasification Technology Comparison (Figure 25)

Parameter	Units	RefCoM	SOLCON	SERI HS
Cost-Optimal HRT	days	23	25.5	40
SRT	days	23	51	67
Ist Order Rate Constant	day^{-1}	0.155	0.155	0.10
Influent TS%	% TS	16.5	8.7	53.6
VS Loading Rate	lbs VS/cu ft/day	0.37	0.15	0.75
	g VS/L/day	5.92	2.40	12.00
VS Conversion %	%	72	82	80
Effluent TS%	% TS	8.0	4.0	27.9
Methane Yield	scf CH_4/lb VS_a	5.03	5.72	5.61
	ml CH_4/g VS_a	314	357	350
SNG Production	10E6 scf/day	2.27	2.58	2.53
	10E3 m^3/day	64.3	73.1	71.6
Active Digester Volume	10E3 cu ft	1,302	3,148	637
	m^3	36,900	89,200	18,000
Total Cost	$/MMBtu gas	$15.09	$11.25	$10.53
% of Cost in Preprocessing		18%	21%	23%
Biogasification		37%	31%	26%
Gas Utilization		9%	12%	10%
Postprocessing		1%	1%	1%
Landfilling		36%	36%	40%
% of Income in Tipping Fees		61%	72%	79%
Recyclables Sale		6%	7%	7%
Gas Sale		33%	21%	14%
Levelized Gas Cost	$/MMBtu	$5.00	$2.37	$1.48

Note: Biogas cleaned up to SNG, all residues landfilled.
Biogasification tipping fee = $40/ton.

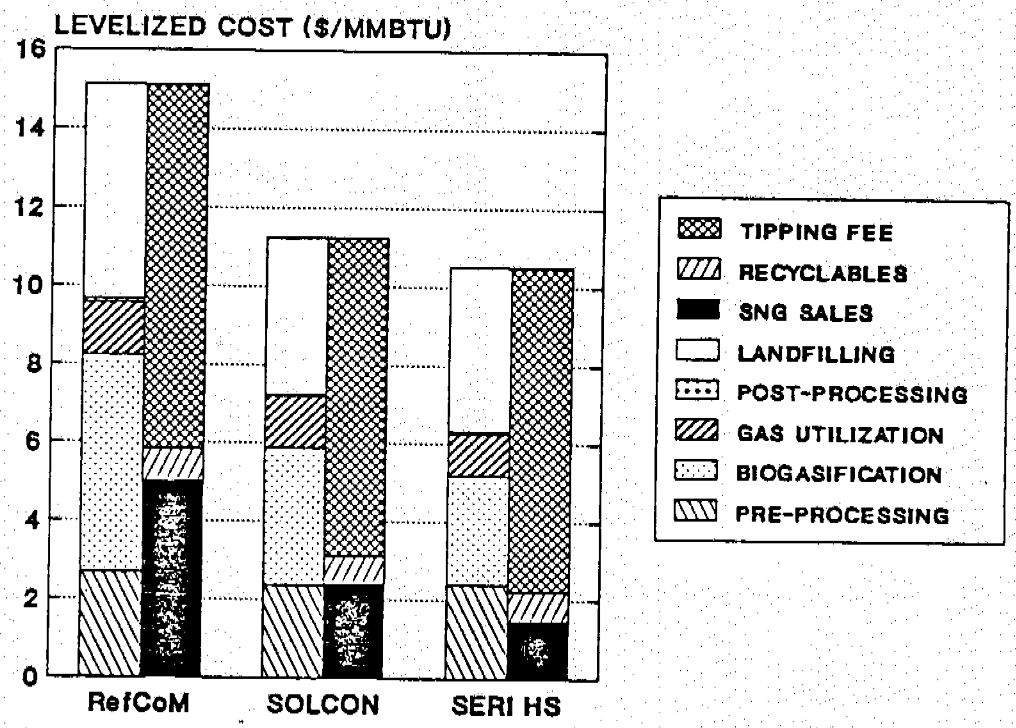


Figure 25: Levelized costs and income streams for three digester technologies. Costs per module in left half of bar diagrams, incomes per source in the right half. The operational parameters for this figure are listed in Table 6.

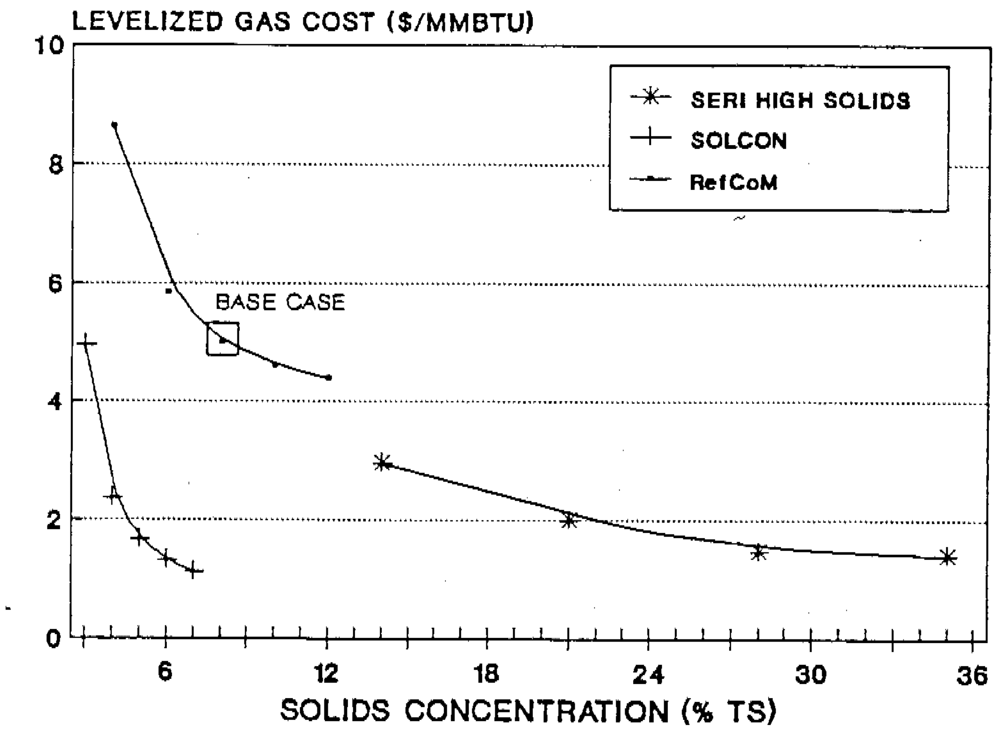


Figure 26: Levelized gas cost versus reactor (= effluent) solids concentration for three digester technologies. Note that RefCoM data above 8% TS are speculative, as are SOLCON data above 4% TS. For SERI HS, a dry matter density of 16.8 lbs TS/cu. ft. (270 g TS/L) was assumed.

SERI High Solids concept is very sound: the high viscosity allows low mixing intensities, and the high TS content maximizes SRT or can minimize reactor volume.

3.4 ENERGY EFFICIENCY SENSITIVITY

3.4.1 Introduction

Anaerobic digestion of MSW could potentially supply the U.S. with 1-2 quad of gas per year (Legrand and Warren, 1986). How much of this energy is actually recovered depends on the efficiency of the process. Anaerobic digestion generates fuel gas but uses process energy, which is almost entirely electricity. Note that at temperatures found in the conterminous U.S., MSW biogasification is thermally self-sufficient once metabolic heat production is included. Consequently, in this discussion no purely thermal process needs are included.

From a societal point of view, it is important to know how much net energy is produced after this process energy has been taken into account. Fuel and electricity are not thermodynamically comparable forms of energy to the extent that electricity is generated from fuel. In the U.S., most electricity is generated from fossil fuel. To achieve an "apples and apples" comparison for this study then, electricity flows were converted to primary fuel equivalents, using an electricity generating efficiency of 30 percent. In other words, 100 Btu (100 J) of fuel generate 30 Btu (30 J) of electricity, for a heat rate of 11,377 Btu/kWh (12.00 MJ/kWh). This convention can also be visualized as follows: assume that a fraction of the biogas was diverted to generate process electricity with an energy efficiency of 30 percent. In this study, the process energy streams are expressed in units of biogas that would have to be diverted to generate the electricity necessary to run the facility.

3.4.2 Process Energy Streams

With this convention in mind, Figure 27 was drawn: the main process energy streams are listed for each of the three technologies considered using the parameters listed in Table 6. RefCoM is distinguished by its very high mixing energy demand (0.53 hp/1000 gal or 105 W m^{-3} at 8% TS). Forty-four percent of the gross biogas energy production would be required to generate the mixing power. SOLCON is unmixed, making it very energy efficient. SERI HS only needs 0.4 hp/1000 gal (79 W m^{-3}) for mixing, and this is applied to a much smaller volume than the other two technologies (see Table 6). Note that the mixing energy requirement for SERI HS can potentially be cut to a fraction of the present value by operating the mixer at less than 1 rpm. It was also assumed that SERI HS produces biogas at 65% CH_4 rather than 55%, which reduces the energy cost of gas cleanup somewhat (for all three technologies, cleanup to SNG is assumed in this figure). Using Figure 27, it is possible to compare the process energy needs of the three processes studied, and to compare these internal energy requirements to the gas production.

3.4.3 Recovery of MSW Energy Content

For a comprehensive energy assessment, it is necessary to include the energy content of the MSW in the analysis, and to examine how efficiently this energy content is recovered. This was done in Figure 28, a diagram of energy flows around the system. All energy flows are compared to the energy content of the incoming MSW which is set at 100 and is estimated by assuming 4,500 Btu/wet pound of MSW (10.46 MJ/wet kg of MSW). The reader is reminded that this analysis focuses on the RefCoM base case (see Tables 1 and 2) with biogas upgrade to SNG and landfilling of all residues.

As can be seen from Figure 28, for every 100 Btu of MSW entering the system, 33 Btu of primary utility fuel are necessary to generate the needed

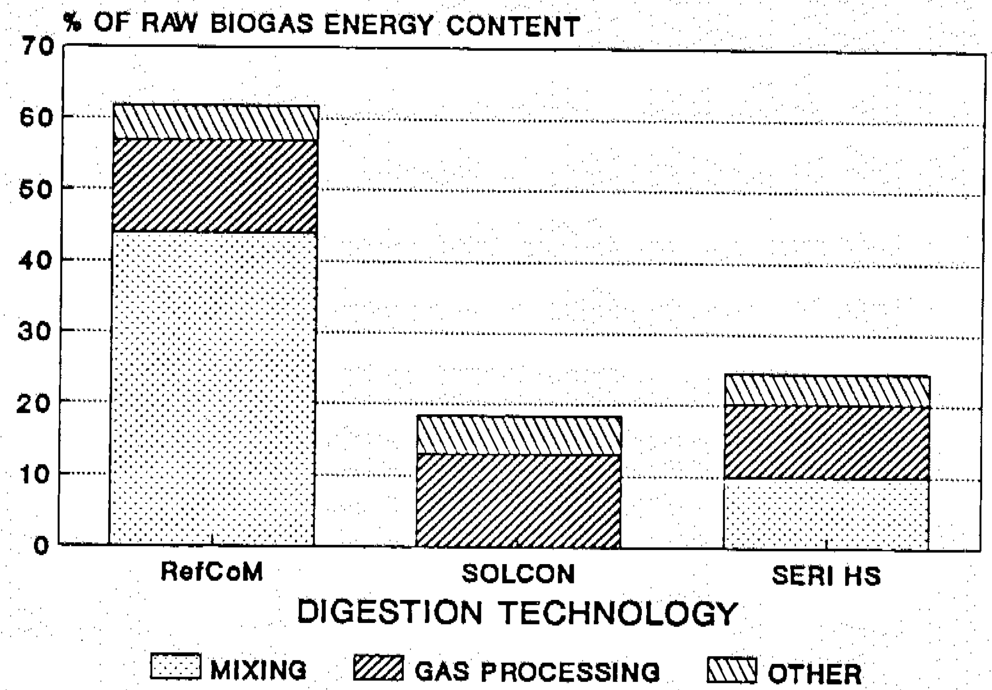


Figure 27: Process energy needs, expressed as percentage of biogas energy content, for three biogasification processes. No net heat addition is contemplated since the processes are thermally self-sufficient. Electricity flows are converted to fossil fuel equivalents using a 30-percent electricity-generating efficiency. Biogas is cleaned up to SNG, all solid residues are landfilled.

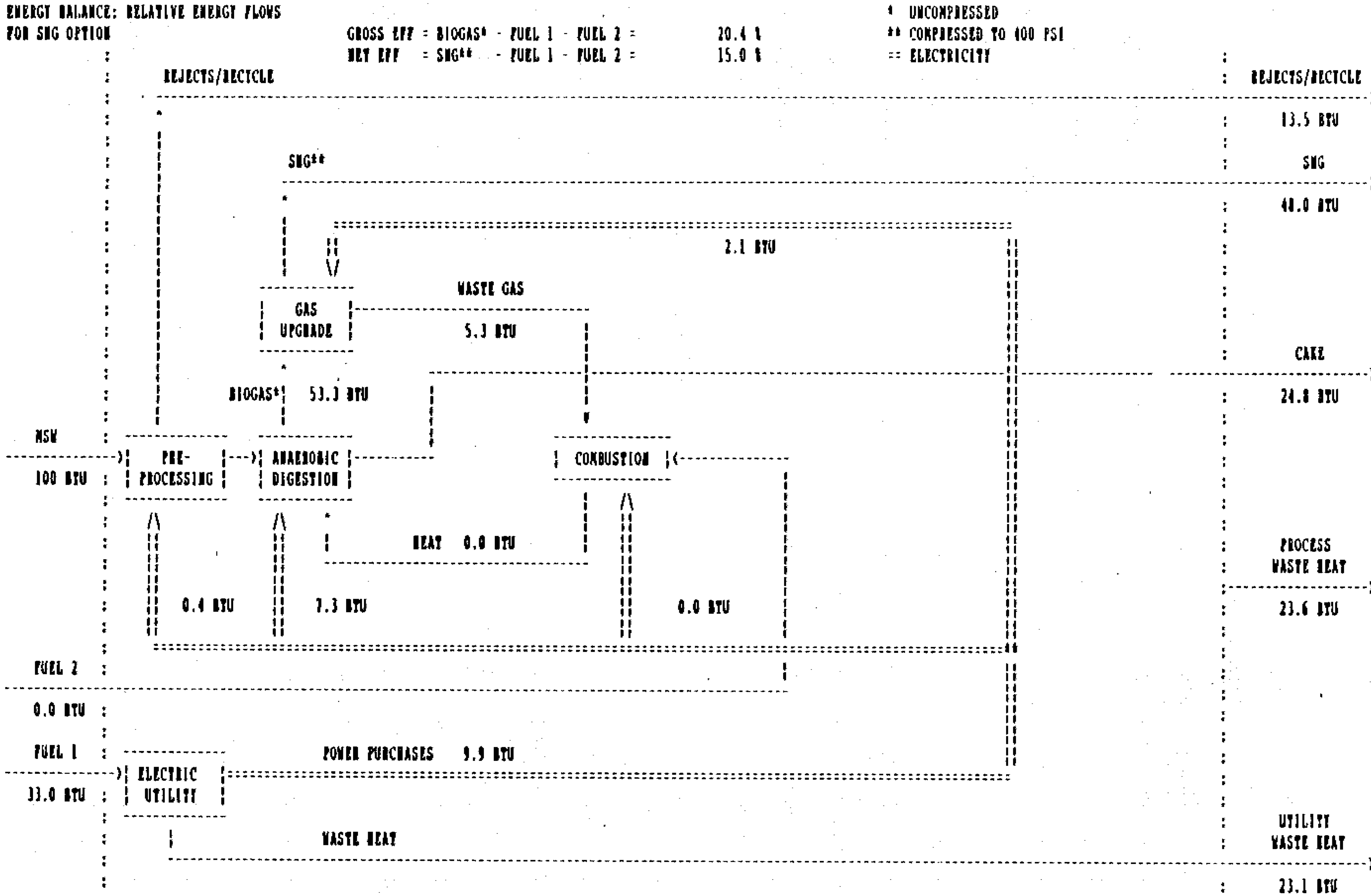


Figure 28: Summary of energy flows, RefCoM base case (biogas upgraded to SNG, all solid residues landfilled). All energy flows are expressed as a percentage of the MSW's energy content. Double lines represent electricity flows, single lines represent fuel value or heat streams. All flows are calculated directly, except the process waste heat which is found by difference.

electricity. Forty-eight Btu of SNG are produced, 23 Btu of heat are lost in the generation of electricity, and 38 Btu is embodied in materials landfilled (rejects and filtercake).

These relationships are displayed in bar graph format in Figure 29. The left half of the bar represents the energy outputs, the right half the inputs. On the left side, the SNG energy can be found at the top, the three types of process power (expressed as utility fuel equivalents, see Section 3.4.1) in the middle, and all losses (heat and residues) at the bottom. On the right side, the MSW and utility fuel inputs are represented. For every 100 Btu of MSW, 48 Btu of SNG are produced, but 33 Btu of utility fuel are needed, therefore, the net energy recovery from MSW is only (48-33) ÷ 100 = 15%. But the process is still a net energy producer since the production of SNG exceeds the supplemental fuel requirement.

3.4.4 Energy Recovery Versus Gas Utilization Choice

So far, the energy recovery from MSW is poor in part because so much of the MSW energy content (38 percent) remains unused, being embodied in the landfilled residues. This can be addressed by coupling anaerobic digestion with combustion of the solid residues and power generation (see Figure 30). This boosts the net energy recovery from 15 to 29 percent. Note that in this RefCoM base case, the power generated from solid residue combustion is insufficient to meet process needs, so some electricity purchase is still necessary. As can be seen from Figure 30, energy losses have been reduced because no combustible material is wasted. The "other" process energy item is slightly increased because of electricity needed to run the power plant (fans, etc.). Note also that the right half of the bar diagram is left out: MSW energy content is always 100 and supplementary utility fuel is equal to the height above 100.

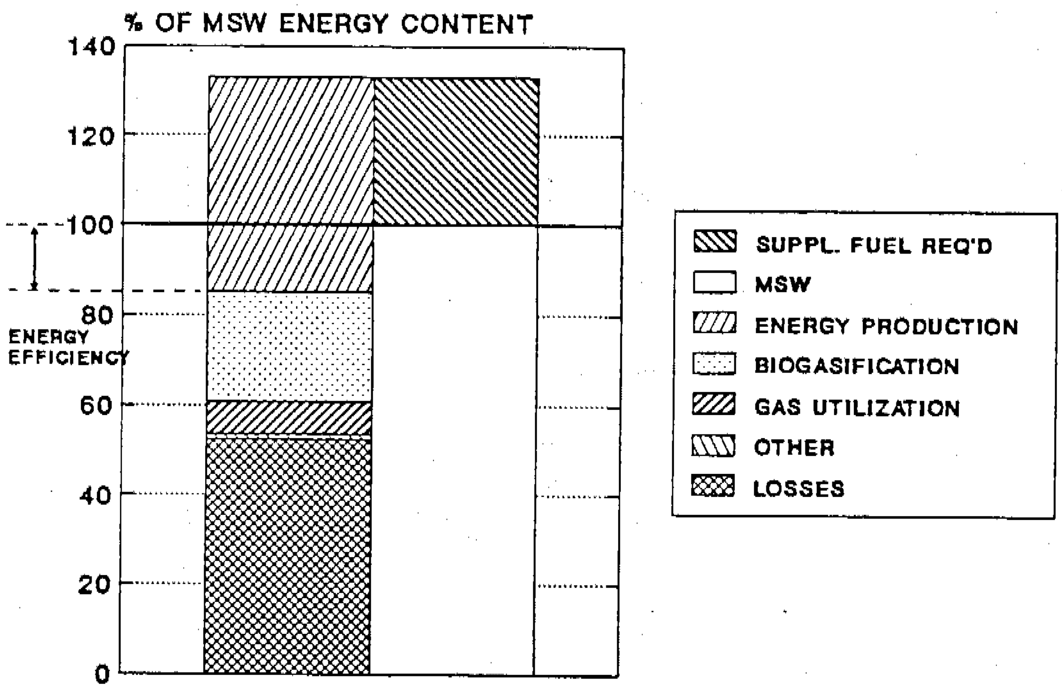


Figure 29: Energy flows expressed as a percentage of the MSW's energy content, RefCoM base case. This diagram is an alternative representation of the data in Figure 28. Left half of bar diagram, top to bottom: energy (SNG) production, three process energy streams, losses (= heat and residue energy content). Right half = MSW energy content (= 100), supplemental fuel necessary to generate process power (= height above 100). Net energy production = portion of energy production below 100 line. Process is a net producer because the energy production is larger than the supplemental fuel required.

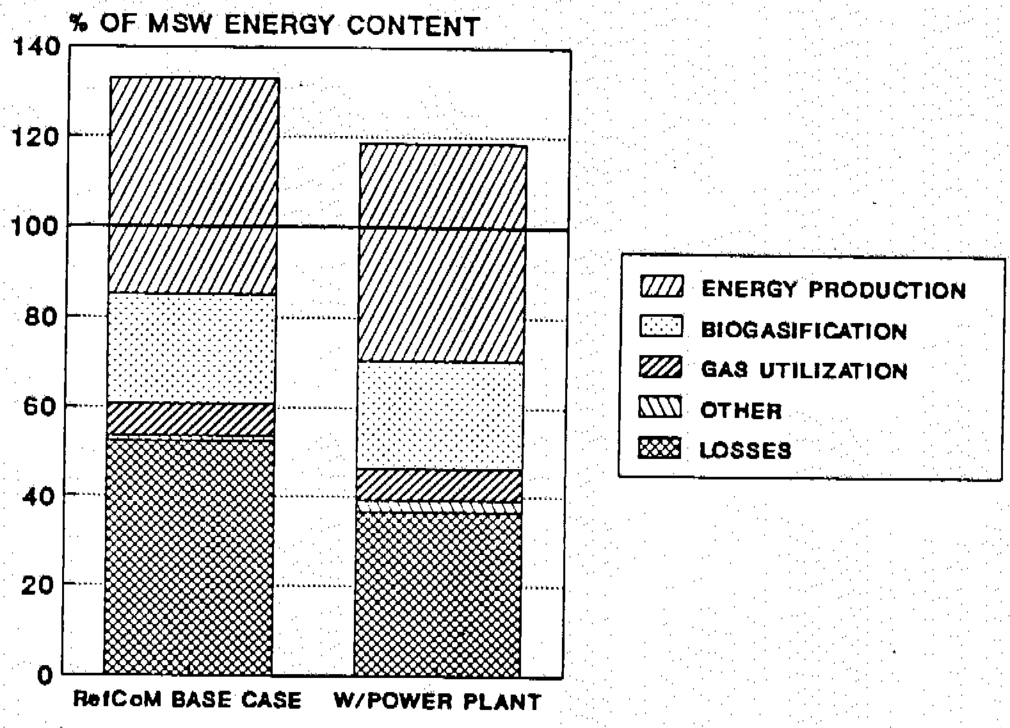


Figure 30: Energy recovery from MSW, RefCoM residues landfilled versus residues burned with power recovery. In both cases the biogas is upgraded to SNG. Format similar to Figure 29, except that right half left out. 100 = energy content of incoming MSW, utility fuel = height above 100, net energy production = energy production below the 100 line.

How gas utilization choices impact the energy balance is explored on Figure 31. Marketing medium Btu gas minimizes the cost of gas upgrading and prevents the gas losses inherent in such cleanup. This can be seen from the left bar in Figure 31: losses are reduced and energy production is increased, each by ten percent compared to the SNG base case, while gas processing energy needs are drastically cut. The resulting energy efficiency is 26 percent, as compared with 15 percent for the SNG base case (second bar from the left). If all biogas is converted to power using a gas turbine, enough power is generated to satisfy the process power needs and no outside power needs to be purchased. The energy efficiency is 29 percent; keep in mind that this is the fossil fuel equivalent of the power generated (see also introduction of this section).

If methanol is produced from biogas (bar on the far right), the process is not a net energy producer, in other words, more outside fuel energy is used than methanol energy is produced. This is due to inefficiencies in the power-hungry methanol manufacture process (see the "losses" segment in the bar diagram). However, this is still an improvement over conventional methanol manufacture from natural gas which has energy efficiencies around 50 to 60 percent. Here, every Btu of fossil fuel used results in the production of 0.84 Btu of methanol (methanol/fossil fuel efficiency = 84 percent). It must be pointed out that using a more energy efficient technology than RefCoM, methanol production from biogas is a net energy producer resulting in methanol/fossil fuel "efficiencies" above 100 percent.

3.4.5 Energy Recovery Potential of the Three Biogasification Technologies

The three technology base cases as summarized in Table 6 are compared for energy efficiency in Figure 32. In all cases, biogas is upgraded to SNG and all residues are landfilled (per base case definition). As noted earlier,

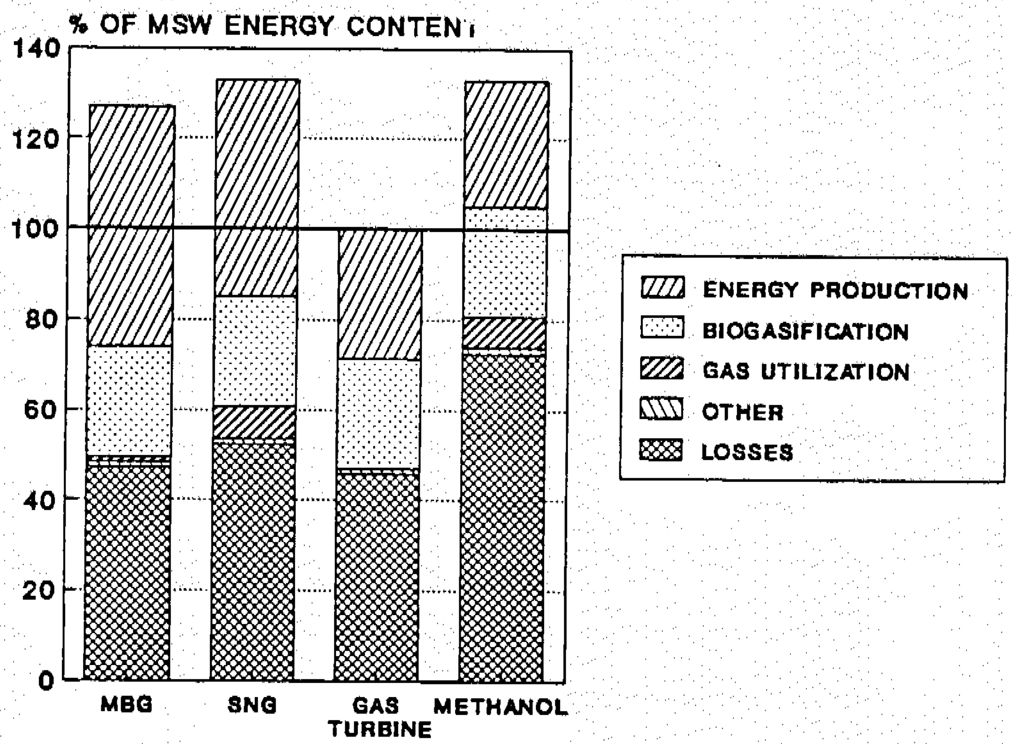


Figure 31: Energy recovery from MSW, RefCoM, four gas utilization options. From left to right the energy products are: medium Btu gas, SNG, electricity (actually the fossil fuel equivalent of that electricity), methanol. 100 = energy content of incoming MSW, utility fuel = height above 100, net energy production = energy production below the 100 line. Note that RefCoM MSW-to-methanol is not a net energy producer.

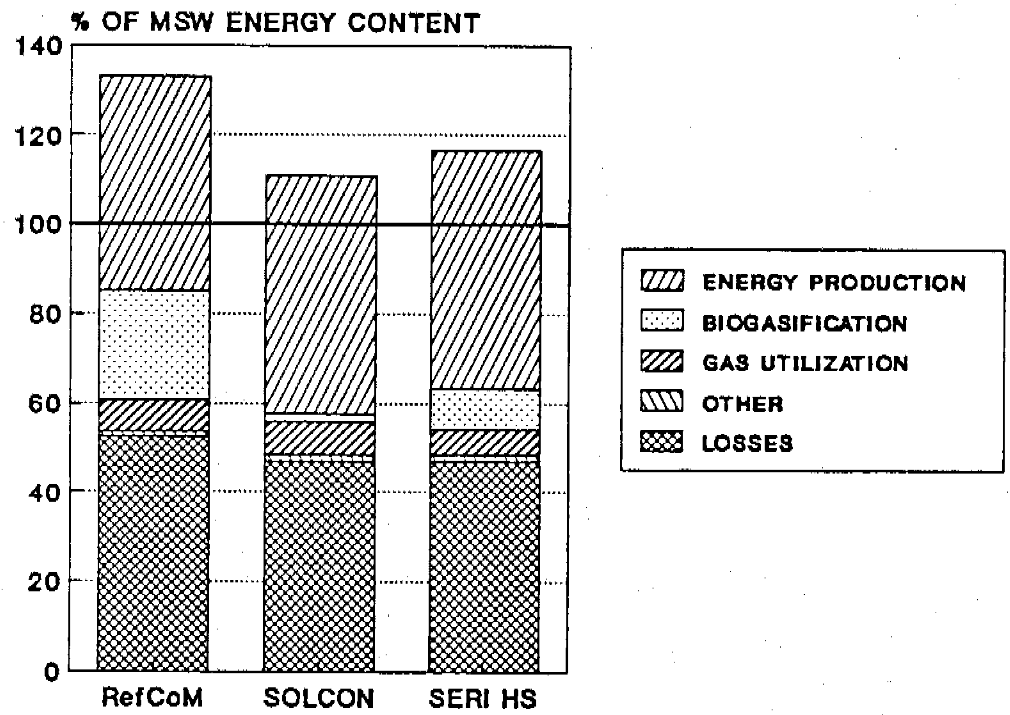


Figure 32: Energy recovery from MSW for three biogasification technological options. Parameters as in Table 6. 100 = energy content of MSW, utility fuel = height above 100, net energy production = energy production below the 100 line. Keep in mind that the cost-optimal RefCoM gas production is a little lower than the two others (see Table 6).

RefCoM is handicapped by a high mixing energy demand and a slightly lower gas production than the two others at its cost-optimal operating point (see Table 6). The efficiencies of energy recovery from MSW are 15 percent for RefCoM, 42 percent for SOLCON, and 37 percent for SERI HS.

These base cases are not optimized for energy efficiency. To illustrate the energy recovery potential of these three technologies, process combinations that maximize energy production were selected, namely medium Btu gas production and combustion of all residues with full power generation. These residues include the front-end plastics-enriched stream and the undigested dewatered effluent or filtercake. The resulting energy balances are illustrated in Figure 33. The efficiency of energy recovery from MSW is 46 percent for RefCoM, 70 percent for SOLCON, and 64 percent for SERI HS. These are not the lowest cost options though, which illustrates that the most economical process is not necessarily the most energy efficient, at least at today's energy prices (assuming $3/MMBtu for gas, 5¢/kWh for electricity purchase and 3¢/kWh for electricity sale). As energy prices rise, however, the more energy-efficient processes will be favored.

3.5 OPTIMIZATION

For each of the digester technologies analyzed--RefCoM, SOLCON, and SERI High Solids--the base cases were optimized by process option. In all instances, recycling of the plastics-enriched stream is economically superior to landfilling of plastics, and composting of the solid residue is more economical than landfilling, incineration, or power generation for the residue. All four of the gas utilization options are viable, depending on the markets for the resultant energy products--medium Btu gas, SNG, electricity, or methanol. An optimum retention time was determined for each digester

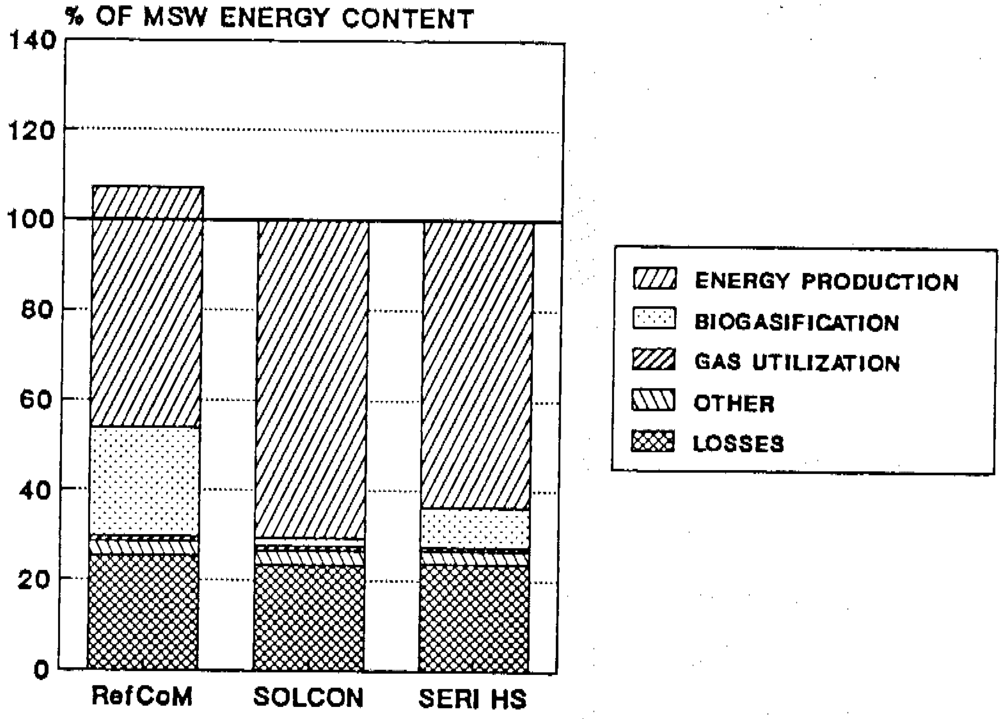


Figure 33: Energy recovery from MSW for three biogasification technological options, energy-optimal process (medium Btu gas produced, all solid residues burned with power generation). 100 = energy content of MSW, utility fuel = height above 100, net energy production = energy production below the 100 line. Only RefCoM requires some outside power purchases.

technology using the process options described above, but with the restriction that the VSCE must be at least 70%; although shorter retention times sometimes produced lower levelized energy product costs, solids destruction is the first priority in MSW biogasification. The following optimal retention times result:

Digester Technology	SRT	HRT
RefCoM	21 days	21 days
SOLCON	25 days	12.5 days
SERI High Solids	32 days	19.1 days

Note that these cost-optimal cases are achievable using current technology, since only the process options and retention times have been modified from the base case for each digester technology. If other parameters such as facility availability, kinetic reaction rate constant, or the value of recyclables were also improved, the results would be even more favorable.

Figures 34 through 37 show the cost-optimal cases for each of the gas utilization options. The reader is reminded that the front-end plastics-enriched stream is disposed of at no cost or income, as is the compost prepared from the digested residue. The retention time is such that at least 70 percent of the volatile solids are converted to gas. The format of Figures 34 through 37 is consistent with previous figures: the levelized costs (preprocessing, biogasification, postprocessing, landfilling) expressed in $/MMBtu are stacked up in the left hand of each bar diagram, whereas the income streams required to pay for these costs are listed in the right half. Key parameters are listed in Table 7.

Marketing of (largely unpressurized) medium Btu gas is considered first, in Figure 34. As can be seen from that figure, a tipping fee of $40/ton and the sale of recovered aluminum at $800/ton already bring in more income than

Table 7: Parameter Values used for 500-tpd MSW Optimized Cases (Figures 34 through 37)

General Assumptions: Plastics-enriched stream recycled at $0, solid residue refined to compost marketed at $0, SRT such that at least 70 percent of VS converted, aluminum sold at $800/ton.

Parameter	Units	RefCoM	SOLCON	SERI HS
HRT	days	21	12.5	19.1
SRT	days	21	25	32
Influent TS%	% TS	16.1	7.35	47.33
VS Loading Rate	lbs VS/cu ft/day	0.39	0.25	1.38
	g VS L^{-1} day^{-1}	6.25	4.00	22.11
% VS Converted	%	70	73	70
Effluent TS%	% TS	8.0	4.0	27.4
Methane Yield	cu ft CH_4/lb VS_a	4.93	5.12	4.91
	ml CH_4/g VS_a	308	320	307
v/v/d Methane		1.92	1.27	6.78
Active Digester Volume	10E3 cu ft	1,224	1,919	345
	m^3	34,600	54,300	9,800

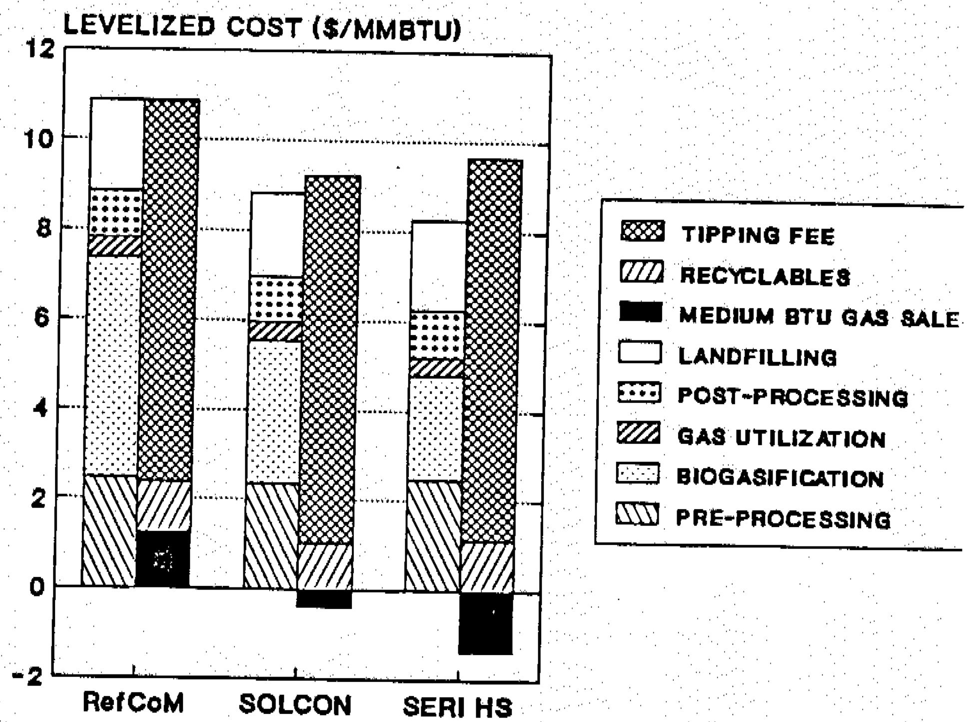


Figure 34: Levelized costs and credits by digester technology for mediur gas production. The MBG must be sold at approximately $1.30/M for RefCoM, but SOLCON and SERI HS generate enough revenues tipping fee and recyclables such that MBG sale is not necess (Tipping fee could be reduced if a market exists for the MBG.)

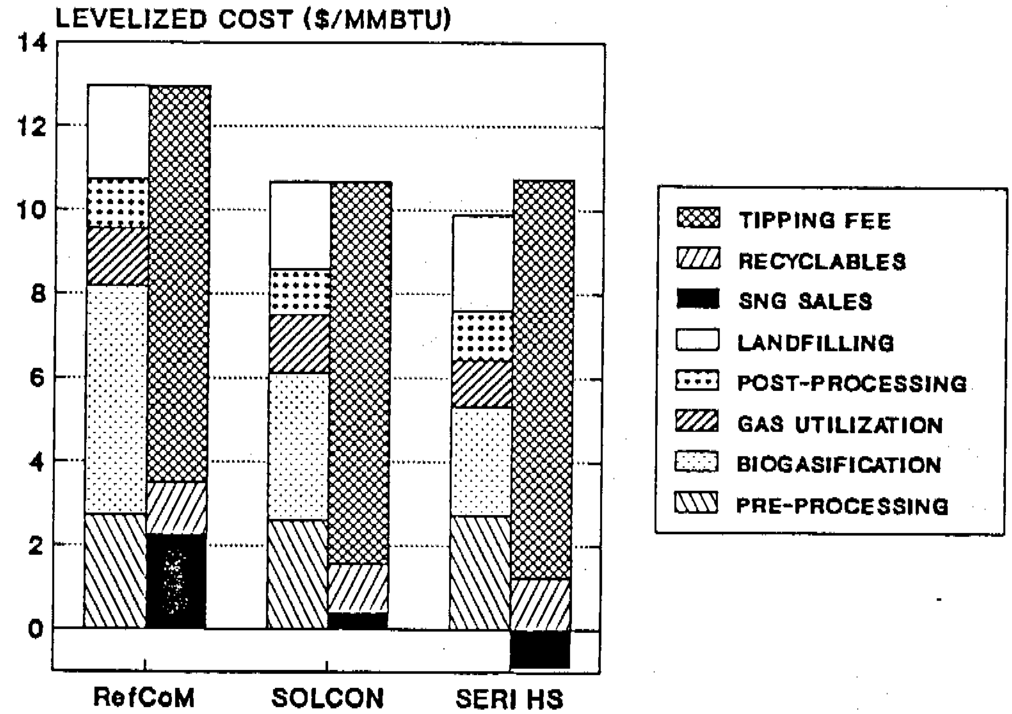


Figure 35: Levelized costs and credits by digester technology for SNG production. The SNG must be sold at a reasonable price for RefCoM and SOLCON facilities to break even, but SERI HS does not require SNG sale at the $40/ton tipping fee.

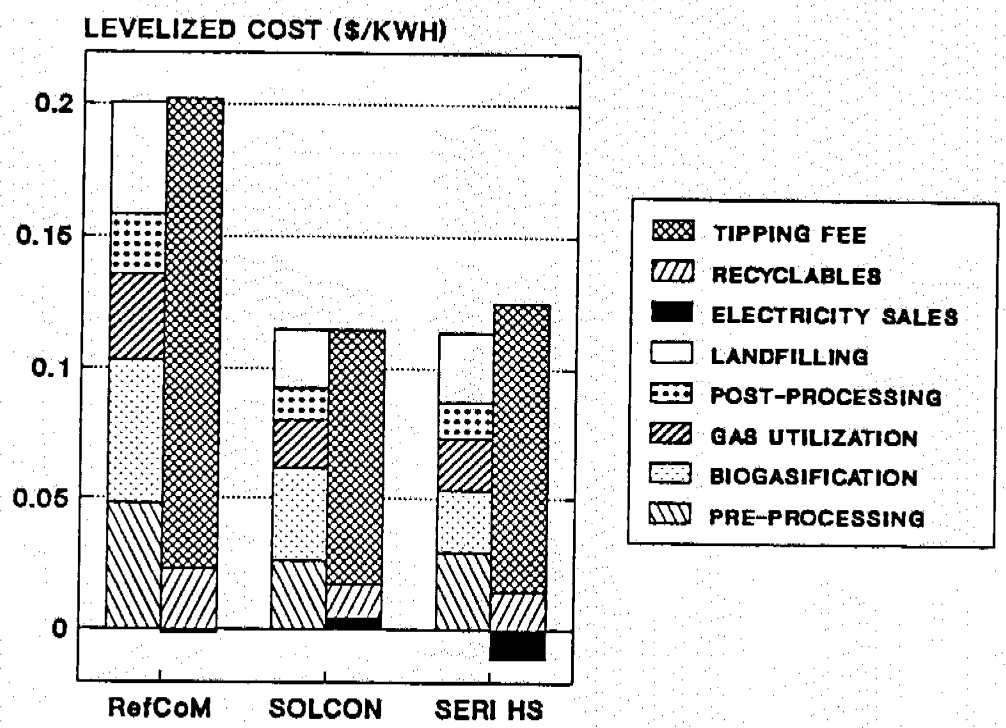


Figure 36: Levelized costs and credits by digester technology for electricity generation by gas turbine. This option is advantageous for all three digester configurations, since SOLCON only requires electricity sale at less than 1¢/kwh, and RefCoM and SERI HS generate enough revenues even without electricity sale.

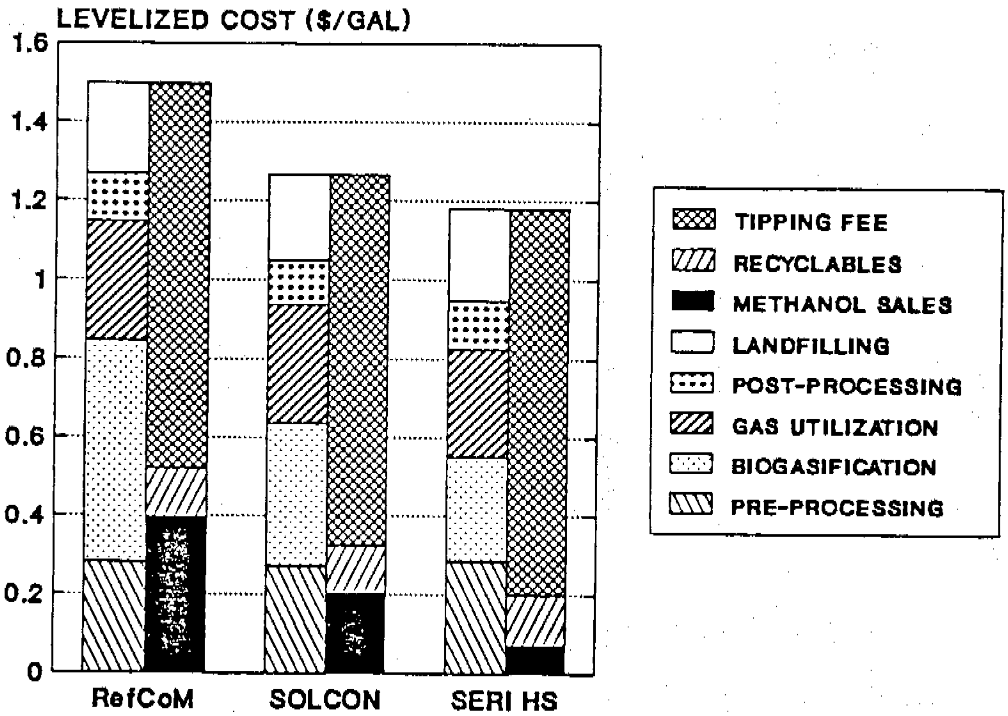


Figure 37: Levelized costs and credits by digester technology for methanol production. These cases all require methanol sales, but at reasonable prices of 40¢/gal for RefCoM, 20¢/gal for SOLCON, and 7¢/gal for SERI HS.

charged are extremely low: 42¢/MMBtu for SNG, less than 1¢/kwh for electricity, and 20¢/gal for methanol. SERI HS facilities generate enough revenue from tipping fees and sale of recyclables to meet expenses without the sale of their energy products, except for methanol which must be sold at $0.07/gal.

In reality, whenever the needed price of an energy product is "negative" or well below market, the tipping fee will be reduced accordingly. Typically, the energy product will be marketed at the going rate and the tipping fee will be calculated. In Table 8, tipping fees are calculated for each type of energy product, assuming a typical U.S. rate for each product. Note that, contrary to the practice so far in this report, the haul and landfill tipping fee is not equal to the biogas facility tipping fee, but was kept constant at $40/ton landfilled. As can be seen from Table 8, RefCoM requires biogasification tipping fees of $32 to $37/ton MSW, SOLCON $23 to $30/ton MSW, and SERI HS $20 to $24/ton MSW. Predictably, MBG marketing consistently leads to the lowest tipping fees since it involves the least amount of gas processing.

These tipping fees compare to the following contemporary values for the U.S. The 1988 Waste Age survey of tipping fees found a U.S. average of $27/ton at landfills and $40/ton at resource recovery (burning) facilities. A more recent survey (BioCycle, March 1990) found landfill tipping fees ranging from $3 to $120/ton and resource recovery tipping fees between $15 and $120/ton. With tipping fees as low as $20/ton, anaerobic digestion is thus certainly a competitive MSW disposal technology, especially using a high solids design such as SERI HS. This is even more evident since the economics of anaerobic digestion are conservatively estimated here with residue tipping fees of $40/ton; no credit being taken for plastics, ferrous metals, or

Table 8: Tipping Fees for Three 500-tpd MSW Anaerobic Digestion Technologies, Four Energy Products.

General Assumptions: Plastics-enriched stream recycled at $0, solid residue refined to compost marketed at $0, SRT such that 70 percent of VS converted, aluminum sold at $800/ton

Parameter	Units	RefCoM	SOLCON	SERI HS
1) MBG Marketing at $3/MMBtu	10E6 scf MBG/day	4.27	4.44	3.60
	10E3 m^3 MBG/day	121	126	102
	MMBtu MBG/day	2,348	2,442	2,340
Tipping Fee	$/ton MSW	31.9	23.4	19.5
2) SNG Marketing at $3/MMBtu	10E6 scf SNG/day	2.23	2.31	2.22
	10E3 m^3 SNG/day	63	65	63
	MMBtu SNG/day	2,118	2,194	2,109
Tipping Fee	$/ton MSW	36.9	28.7	23.7
3) Electricity* Sold at 3¢/kWh	MWh/day	112	205	181
	MMBtu elec./day	382	690	618
Tipping Fee	$/ton MSW	33.0	29.6	24.3
4) Methanol Sold at 50¢/gal.	gallons methanol/day	20,500	21,300	20,400
	liters methanol/day	77,600	80,600	77,200
	MMBtu methanol/day	1,228	1,276	1,223
Tipping Fee	$/ton MSW	35.7	27.4	22.5

*Generated with biogas-powered gas turbine.

compost (they are given away); and glass being landfilled. We feel this conservatism is justified because of the difficulty of marketing recyclable materials, as illustrated by the used newsprint glut. The recycling picture i· changing rapidly though, as a larger fraction of the waste stream is recycled nationwide. On the other hand, source separation or curbside recycling could remove most recyclables from the waste stream before they reach the biogasification plant.

The potential impact of recyclables and landfill tipping fees other than $40/ton on anaerobic digestion economics can be graphically estimated from Figures 34 through 37. The impact of various recyclable materials prices was illustrated on Figure 22 in Section 3.2.6.

- The dry matter density (grams TS/L) should be measured, preferably at different TS% and depths. The lower limit of TS% for SERI HS operation should be defined. The biologically optimal TS% (with regard to reaction rate, extent of conversion, and stability of operation) should be defined. Taken together, this information would allow a better definition of the optimal TS% and the applicable range for SERI HS.
- SRT should be measured with a tracer test.
- A preliminary full-scale design should make it possible to identify further scale up problems.

3. <u>General Topics</u>:

- The metabolic heat production that accompanies biogasification should be estimated in theory, then demonstrated in a carefully controlled experiment. Some estimates indicate that large-scale MSW digesters may show a heat surplus. That this has not been observed may have to do with the fact that nobody has biogasified a dry feedstock in a full-scale system yet.
- The reliability and sturdiness of MSW anaerobic digestion should be better demonstrated by subjecting a reactor to various upsetting conditions (toxics, temperature shock, etc.).
- The materials handling characteristics (viscosity, etc.) of the digesting slurry as well as the feed material should be measured. This information is necessary to design the materials handling devices (pumps, conveyors, etc.).

4.3 GAS PROCESSING

- Because it has the potential of saving energy and cost, MED should be further developed. A design-oriented model is needed, based on sound design and costing methods. The carrier liquid needs to be easy to handle (little suspended matter) and it must be possible to circulate it through the digester without disturbing the process. The SEBAC (Sequenced Batch Anaerobic Composting) process under development at the University of Florida would be particularly appropriate in this respect. In this process, leachate is run through batches of RDF. A way needs to be found to apply MED to the SERI HS process.

4.4 POST-PROCESSING OF RESIDUES

- Anaerobic digestion can be seen as pretreatment of MSW prior to landfilling since it organically stabilizes MSW which should preclude the formation of landfill gas or strong leachates. This needs to be demonstrated with a side-by-side experiment with raw MSW and digested residue.
- The uncertainty regarding environmental impact of MSW-derived compost should be narrowed by investigating its quality (i.e. degree of stabilization, heavy metals, organic toxics, human pathogens, weed seeds, etc.) and demonstrating its application. Aerobic and anaerobic composts need to be compared.

4.5 PRODUCT MARKETING

The feasibility of marketing MBG, SNG, methanol, and compost should be studied.

4.6 SYSTEMS ANALYSIS

- The societal benefit from extending landfill life needs to be quantified, in $/ton.
- With three plastics processing options, three to five biogasification technologies, four to five gas processing options, four residue disposal options, and five to ten parameters to analyze, up to 3,000 sensitivity analyses can be performed. Only a tiny, but hopefully representative fraction of all possible analyses was attempted here. As processes such as MED or SERI HS become better defined, new analyses will be needed. One could, for example, examine the implications of a base case comprising SERI HS, MED, and methanol production.

4.7 PRIORITIES

The highest priority should be given to the following research:

- Pilot-scale operation of the SERI HS process, optimized with regards to TS%, temperature, and mixing rpm.
- MED pilot operation using SEBAC, or if possible SERI HS.
- Environmental and marketing aspects of refining residue to compost.

INPUT SECTION

Tons per day MSW (7 days/wk)	500 tpd
% Total Solids MSW	74.00 %
% Volatile Solids (of TS) MSW	73.00 %
% Biodegradable Volatile Solids (of VS) MSW	87.00 %
MSW Tipping Fee (1990 $)	$40.00 /[illegible]
SRT (Solids Retention Time)	23.00 days
SRT/HRT ratio (CSTR if 1)	1.00
Treatment of plastics option:	1
1 = Landfill	
2 = Recycle (no value or cost)	
3 = Burn (only if incineration or power plant option is chosen)	
Biogas processing option:	2
1 = Sale of medium-Btu gas	
2 = Gas cleanup to SNG	
3 = Gas turbine	
4 = Methanol process	
5 = MED process for high Btu gas (ITERATIVE CALC)	
Solid residue processing option:	1
1 = Landfill	
2 = Incinerate (recover process heat)	
3 = Power plant (recover process heat & generate electricity)	
4 = Landspreading / compost sales	
Low, average or high cost estimate (L, A, H)	A
Base year dollars (1980 - 1991)	1990

OUTPUT SECTION

Net daily SNG flow	2.27 MMscf/day
Net SNG production	0.67 10E12 Btu/yr
Volatile Solids Conversion Efficiency %	71.76 %
Gross methane yield	5.03 scf CH4/lb VSa
vol. methane/vol. reactor/day	1.84 v/v/d
Volatile Solids Loading Rate (VSLR)	0.37 lb VS/cf-day

LEVELIZED COST	1990 $/MMBtu	% of Total Cost
Pre-processing	$2.6763	17.74%
Biogasification	$5.5318	36.66%
Gas utilization	$1.3688	9.07%
Post-processing of residues	$0.0826	0.55%
Landfilling	$5.4310	35.99%
TOTAL COST	$15.0904	100.00%
MSW Disposal Credit	($9.2571)	-61.34%
Byproduct Credit	($0.8331)	-5.52%
TOTAL CREDIT	($10.0902)	-66.86%
NET TOTAL	$5.0002 /MMBtu	33.14%

MASS / ENERGY BALANCE ASSUMPTIONS

PRE-PROCESSING	
TS% into wet trommel	60.00 % TS
Density of MSW	37 wet lb/cf
Density of refuse separation rejects	60 wet lb/cf
Density of plastics	20 wet lb/cf
Aluminum content of MSW	0.45 %
CONVERSION	
Percent methane in biogas	55 %
Digester temperature	140 deg F
TS% digester effluent	7.95 % TS
TS% filtercake	50 % TS
TS% recycled filtrate	2.0 % TS
Dry matter density of feed	12 lbs TS/cf
Max. digester height	50 ft
Max. digester diameter	50 ft
Min. number of digesters	2
Average ambient temperature	55 deg F
Design ambient temperature	0 deg F
Insulation thickness	2 in
Density of filtercake	60 wet lb/cf
MED MODULE	
Digester-to-stripper volume ratio	20
Liquid recirculation rate	1.5 v/v/d
Air-to-water ratio	30
BOILER/GENERATOR	
Boiler efficiency	70.00 %
TS% in ash from boiler	70 % TS
Dry, ash-free residue heat content	10,000 Btu/DAF lb
Density of combined ash	74 wet lb/cf
COMPOSTING	
Percent oversize removed by trommel	25 %
Specific gravity of raw compost	0.5
Volume per windrow	1333 cf
Storage period	7 days
Floor area per windrow	1400 sf
Density of trommel rejects	20 wet lb/cf
ENERGY BALANCE	
Utility generating efficiency	30 %
MSW heat content	4,500 Btu/lb
Mixing energy (ENTER NA FOR CSTR CALCULATION)	NA kwh/day

FEED INPUT MODULE

Q = Wet Tons Feed / day
TS% = Total Solids Content (dry weight %)
VS% = Volatile Solids Content (dry ash-free weight %), as a % of TS
BVS = Biodegradable Volatile Solids
TVS = Total Volatile Solids
k = 1st order reaction rate coefficient (1/day)
COD/VS = lbs of chemical oxygen demand per lb of VS
VSCE = VS Conversion Efficiency expected

		Q	QxTS	QxTSxVS	BVSCE	VSCE	QxTSxVS xVSCE	METHANE SCF/DAY	BVS
Stream 1 :RDF									
Q	443 tpd								
TS%	62.28 %								
VS%	86.44 %	443	275.76	238.36	0.78	0.72	171.05	2,398,987	219.03
(BVS/TVS)%	91.89 %								
k	0.155 (1/day)								
COD/VS	1.25								
Stream 2 :									
Q	0 tpd								
TS%	0.00 %								
VS%	0.00 %	0	0.00	0.00	0.00	0.00	0.00	0	0.00
(BVS/TVS)%	0.00 %								
k	0.00 (1/day)								
COD/VS	0.00								
Stream 3 :									
Q	0 tpd								
TS%	0.00 %								
VS%	0.00 %	0	0.00	0.00	0.00	0.00	0.00	0	0.00
(BVS/TVS)%	0.00 %								
k	0.00 (1/day)								
COD/VS	0.00								

	Q	QxTS	QxTSxVS	BVS/TVS (AVG %)	VSCE (AVG)	QxTSxVS xVSCE	METHANE SCF/DAY	BVS
TOTALS	443	275.76	238.36	91.89	0.72	171.05	2,398,987	219.03

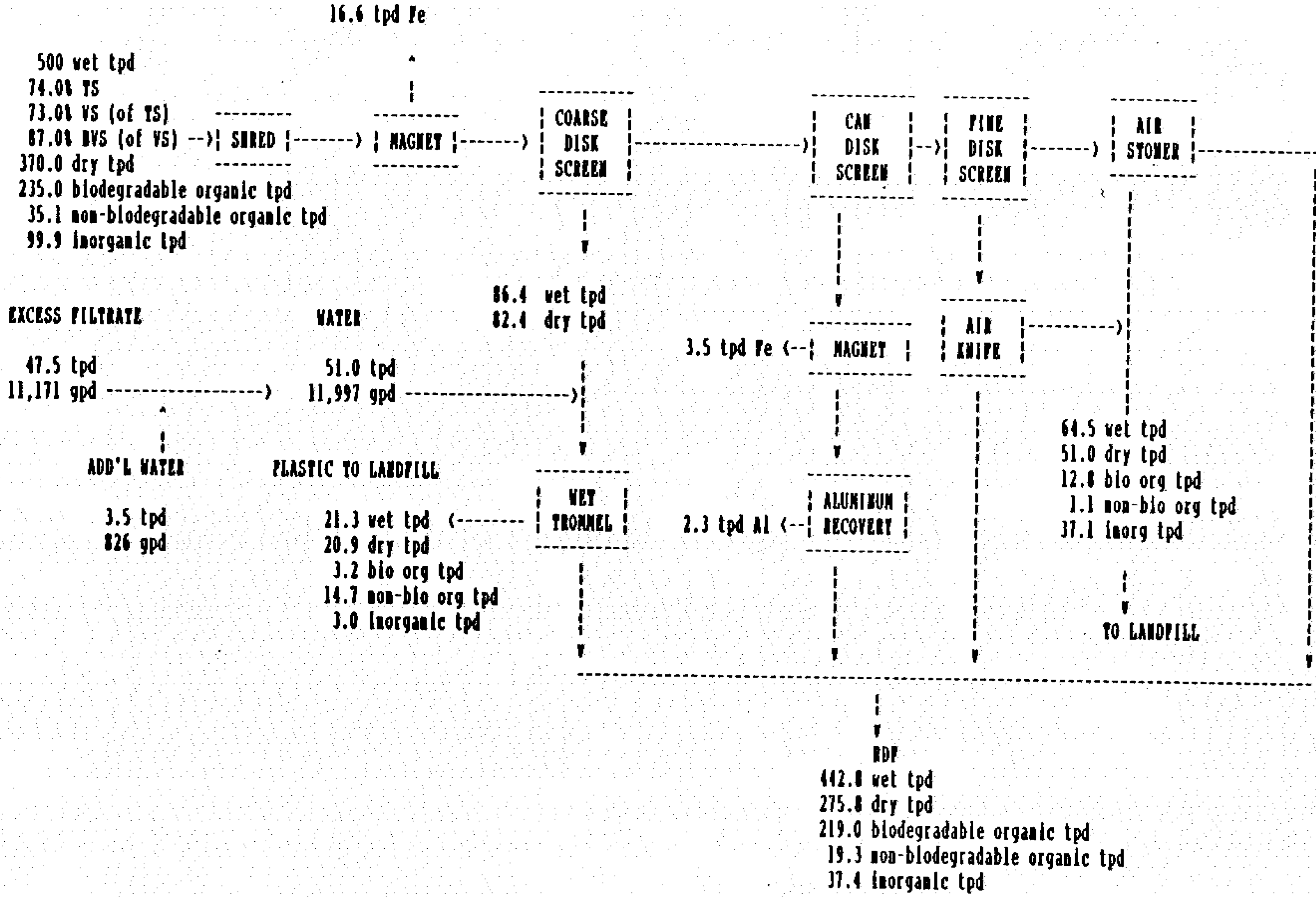
UNIT COST/CREDIT ASSUMPTIONS (1990 $)

Value of scrap iron	$0 /ton
Value of scrap aluminum	$800 /ton
Value of compost	$5 /ton
Value of electricity	$0.030 /kwh
Cost of electricity	$0.050 /kwh
Cost of land	$10,000 /acre
Cost of fuel	$4.00 /MMBtu
Bond financing % of capital	40 %

LEVELIZED COST MODULE INPUTS

to	initial year of plant operation	1990
BLA	book life A in years (long term equipment)	30
BLB	book life B in years (short term equipment)	10
CP	construction period in years	1.5
dc	constant-dollar discount rate	0.05
inf	inflation rate	0.05
eb	current $ escalation rate of byproduct value	0.05
ew	current $ escal. rate of waste disposal credit	0.05
ec	current $ construction cost escalation rate	0.05
ef	current $ fuel cost escalation rate	0.05
el	current $ land cost escalation rate	0.05
em	current $ variable O&M cost escalation rate	0.05
fd	fraction financed by debt	1.00
fce	fraction financed by common equity	0.00
fpe	fraction financed by preferred equity	0.00
fnb	fraction financed by nonborrowed funds	0.00
rd	current-dollar return to debt	0.081
rce	current-dollar return to common equity	0.000
rpe	current-dollar return to preferred equity	0.000
rnb	current-dollar return to nonborrowed funds	0.113
Tax Rate	combined state and federal income tax rate	0
TLA	tax life A in years (long term equipment)	15
TLB	tax life B in years (short term equipment)	5
ITC	investment tax credit	0.1
WCF	working capital fraction	0.125
SF	service factor	0.85
PDARS	PDA for refuse separation module	0
PDAC	PDA for conversion module	0
PDAGC	PDA for gas cleanup module	0
PDASRP	PDA for solid residue processing module	0

REFUSE SEPARATION MODULE



CONVERSION

0.88 10E12 Btu/yr
876.21 MMscf methane/year
5.03 scf methane/lb VSa
4,361,795 scf biogas/day
2,398,987 scf methane/day
55 % methane
171.1 dry tpd converted

17.4 tpd evaporated

FEED BLENDING

FEED MIX
WITHOUT RECYCLED FILTRATE

442.8 tpd wet
275.8 tpd dry
62.28 TS%
86.44 VS% (of TS)
91.89 BVS/TVS%
0.6803 lb wet/cu ft-day

--->

REACTOR INFLUENT

1,840 tpd wet
304 tpd dry
432,834 gpd
16.51 TS%

--->

DIGESTERS (CSTR)

TEMP	140 deg F
VSCE	71.76 %
SLURRY VOLUME	
	9.739 10E6 gal
	1,301,833 cu ft
NUMBER	15 digesters**
VSLR	0.37 lbs VS/cu ft-day
HRT	23 days

** 96,432 cu ft each
49.7 ft diameter
49.7 ft height

--->

REACTOR EFFLUENT

1,651 tpd wet
133 tpd dry
388,484 gpd
8.0 TS%

---> DEWATERING --->

FILTERCAKE

208 tpd wet
104 tpd dry
67 tpd organic
37 tpd ash
50.0 TS%

(to solid residue processing)

FILTRATE

1,444 tpd wet
29 tpd dry
339,649 gpd
2.0 TS%

EXCESS FILTRATE

47 tpd wet
1 tpd dry
11,171 gpd
2.00 TS%

(to wet trommel)

RECYCLED FILTRATE

1,397 tpd wet
28 tpd dry
328,649 gpd
2.00 TS%

DILUTION WATER REQUIRED

0 tpd
0 gpd

GAS CLEANUP MODULE

BIOGAS
PRODUCED

4.36 MMscf/day
2,399 MMBtu/day ------------>
55.00 % methane

ELECTRIC COMPRESSORS

320 bhp/MMscfd biogas
1,396 compressor hp
91% motor efficiency
1,144 kw
27,461 kwh/day

PRISM
SEPARATOR
SYSTEM
(INCLUDING
RECYCLE)

------> SYNTHETIC NATURAL GAS
95% METHANE CONTENT

2.27 MMscf/day
2,159 MMBtu/day
90.00 % Methane
recovery

PERMEATE
SIDESTREAM

2.09 MMscf/day
240 MMBtu/day
11.48 % Methane in stream

CONVERSION/BIOGAS PROCESSING SUMMARY

MSW wet tpd	500.0 tpd
Feed dry tpd	275.8 tpd
Feed TS%	62.28 % TS
Slurry volume	1,301,833 cu ft
No. of digesters	15
Usable volume per digester	86,789 cu ft
Aboveground surface area/digester	9,700 sq ft
Actual reactor diameter	49.7 ft
Actual reactor height	49.7 ft
Effluent wet tpd	1651.1 tpd
Effluent dry tpd	132.6 tpd
Filtercake (compost) wet tpd	207.6 tpd
Filtercake organic tpd	67.3 tpd
Filtercake TS%	50.0 % TS
Boiler fuel wet tpd	0.0 wet tpd
Boiler fuel Btu/lb	0 Btu/lb
Gross biogas production	0.88 TBtu/yr
Biogas flow	4.36 MMscf/day
Biogas quality	55.0 % methane
Net SNG production (no downtime)	0.79 TBtu/yr
Net SNG production	0.67 TBtu/yr
SNG flow	2.27 MMscf/day

PROCESS ENERGY MODULE

ELECTRICITY CONSUMPTION

	KWH/DAY		% OF GROSS PROD.
RDF module/front end power	5,000		0.7%
Reactor mixing	92,649		13.2%
Conversi.` process pumping	1,898		0.3%
Mechanical drying of effluent	1,910		0.3%
Boiler plant fans and pumps	0		0.0%
Delumper for composting	0		0.0%
Trommel for composting	0		0.0%
Compression of MBG	0		0.0%
Gas cleanup for SNG option	27,461		3.9%
Gas cleanup for methanol option	0		0.0%
Methanol conversion	0		0.0%
Pumping for MED process	0		0.0%
Air blower for MED process	0		0.0%
Building utilities	1500		0.2%
Total electricity required	130,419	kwh/day	18.5%
Total power required	5.43	MW	
Fuel equivalent of power required	61.82	MMBtu/hr	61.8%

DIGESTION THERMAL BALANCE

	AVG MMBTU/HR	DESIGN MMBTU/HR	% OF GROSS PROD. AVG	% OF GROSS PROD. DESIGN
Feed enthalpy	1.14	0.00	1.1%	0.0%
Dilution H2O enthalpy	0.00	0.00	0.0%	0.0%
Mixing heat input	13.18	13.18	13.2%	13.2%
Metabolic heat input	8.50	8.50	8.5%	8.5%
Wet biogas enthalpy	-0.73	-0.73	-0.7%	-0.7%
Filtercake enthalpy	-1.57	-1.57	-1.6%	-1.6%
Excess filt. enthalpy	-0.55	-0.55	-0.6%	-0.6%
Conduction heat loss	-0.93	-1.54	-0.9%	-1.5%
Recycle heat loss	-1.16	-1.16	-1.2%	-1.2%
Evaporative heat loss	-1.47	-1.47	-1.5%	-1.5%
Net heat req'd (prod)	-16.39	-14.64	-16.4%	-14.6%

TOTAL PRIMARY ENERGY REQUIRED	61.8%	61.8%

COST TABLE (JUNE 1987 $)						CAPITAL COSTS INCLUDING FINANCING	
	LONG-TERM CAPITAL ($)	SHORT-TERM CAPITAL ($)	OTHER O&M ($/YR)	LABOR ($/YR)	FUEL/POWER ($/YR)	LONG-TERM CAPITAL($)	SHORT-TERM CAPITAL ($)
1. REFUSE SEPARATION	4,088,750	4,088,750	171,060	414,450	107,901	5,724,250	5,724,250
2. CONVERSION							
Digesters	5,063,117						
Heating	0				0		
Dewatering		1,989,681			31,708		
Mixing	2,473,482				1,537,995		
Insulation	489,172						
Pumping		364,445			31,503		
Miscellaneous	4,289,193						
Total	12,314,964	2,354,126	256,467	240,920	1,601,206	17,240,950	3,295,776
3. BIOGAS PROCESSING							
Compression for MBG	0		0	0	0		
Gas cleanup for SNG	3,059,528		61,083	87,700	455,865		
Gas utility intertie	38,000						
Gas turbine	0		0				
Gas cleanup for methanol	0		0	0	0		
Package methanol plant	0		0	0	0		
Stripping tower for MED	0		0				
Pumping for MED		0	0		0		
Air compression for MED		0	0		0		
Total	3,097,528	0	61,083	87,700	455,865	4,336,539	0
4. SOLID RESIDUE PROCESSING							
Electric intertie	232,209						
Landfilling			3,311,263				
Boiler	0		0	0			
Boiler and turbine	0		0	0			
Delumper		0	0		0		
Trommel		0	0		0		
Windrow composting		0		0			
Total	232,209	0	3,311,263	0	0	325,093	0
5. GRAND TOTAL	$19,733,451	$6,442,876	$3,799,874	$743,070	$2,164,972	$27,626,831	$9,020,026
TOTAL CONSTR. COST:	$26,176,327						

CREDIT TABLE (1987 $/YR)

1. BYPRODUCT CREDIT		
Scrap iron	$0	/year
Scrap aluminum	$507,965	/year
CO_2 recovery		/year
Excess elec. generated	$0	/year
Compost	$0	/year
Total	$507,965	/year
2. MSW DISP. CREDIT (TIP FEE)	$5,644,060	/year
3. GRAND TOTAL CREDITS	$6,152,025	/year

LEVELIZED COST MODULE

Intermediate Calculations

A	number of acres of land	10
DC	design capacity (MMBtu/yr)	788,586
L	land cost per acre (in tb dollars)	10,000
tb	base year for cost estimation	1987
a	allowed AFUDC rate	0.081
d	current-dollar discount rate	0.103
fe	fraction financed by equity	0.000
re	current-dollar return to equity	0.113
ra	weighted-average after-tax cost of capital	0.081
rm	current-dollar return to O&M cost	0.050
rf	current-dollar return to fuel cost	0.050
rb	current-dollar return to byproduct value	0.050
rw	current-dollar return to waste disposal credit	0.050
rl	current-dollar return to land cost	0.050
TC	tax credit	0.000
LR	land rent	1025

CCR CALCULATION

PART A		PART B	
A	0.063	A	0.126
B	0.940	B	0.940
C	0.064	C	0.126
D	0.000	D	0.000
E	0.466	E	0.621
CCRA	0.051	CCRB	0.109

Cost Inputs: ($1,000,000)

	Refuse Separation	Conversion	Gas Cleanup	Solid Residue Processing	Total System
Capital Pt. A	5.724	17.241	4.337	0.325	27.627
Capital Pt. B	5.724	3.296	0.000	0.000	9.020
O & M - A	0.000	0.418	0.149	3.311	3.878
O & M - B	0.586	0.080	0.000	0.000	0.665
Fuel A	0.000	1.344	0.456	0.000	1.800
Fuel B	0.108	0.257	0.000	0.000	0.365
Land Rent	0.010	0.000	0.000	0.000	0.010
Byproduct Credit	0.508	0.000	0.000	0.000	0.508
MSW Disposal Credit	0.000	5.644	0.000	0.000	5.644
Total	11.644	16.991	4.941	3.636	37.213

Levelized Cost: ($/MMBtu)

	Refuse Separation	Conversion	Gas Cleanup	Solid Residue Processing	Total System
Capital Pt. A	0.436	1.314	0.330	0.025	2.105
Capital Pt. B	0.932	0.536	0.000	0.000	1.468
O & M - A	0.000	0.623	0.222	4.940	5.785
O & M - B	0.874	0.119	0.000	0.000	0.993
Fuel A	0.000	2.005	0.680	0.000	2.686
Fuel B	0.161	0.383	0.000	0.000	0.544
Land Rent	0.015	0.000	0.000	0.000	0.015
Byproduct Credit	0.758	0.000	0.000	0.000	0.758
MSW Disposal Credit	0.000	8.420	0.000	0.000	8.420
Working Capital	0.017	0.051	0.013	0.050	0.130
PDA	0.000	0.000	0.000	0.000	0.000
TOTAL	1.677	-3.389	1.245	5.015	4.548

Levelized Cost Summary: ($/MMBtu)

	Refuse Separation	Conversion	Gas Cleanup	Solid Residue Processing	Landfill	Total System
Capital Pt.	1.368	1.850	0.330	0.025		3.573
O & M	0.874	0.742	0.222	0.000	4.940	6.778
Fuel	0.161	2.389	0.680	0.000		3.230
Land Rent	0.015	0.000	0.000	0.000		0.015
Byproduct Credit	0.758	0.000	0.000	0.000		0.758
MSW Disposal Credit	0.000	8.420	0.000	0.000		8.420
Working Capital	0.017	0.051	0.013	0.050		0.130
PDA	0.000	0.000	0.000	0.000		0.000
TOTAL COST	2.434	5.032	1.245	0.075		13.726
TOTAL CREDIT	0.758	8.420	0.000	0.000		9.178

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June 13, 1990

Robert Legrand
Hunter Services, Inc.
6737 Southpoint Drive South
Jacksonville, FL 32216

Dear Bob,

I have had the draft final report from Hunter Services, Inc., reviewed by several people at SERI and would like to provide a brief summary of their comments to you for incorporation into the final report. Overall, the report was favorably reviewed and perceived to represent a major advance in the understanding of the economic sensitivities of the anaerobic digestion process as it pertains to MSW. Specific comments include:

1 • This report does not provide process flow sheets with appropriate mass and energy balances, equipment lists, and equipment costs. Moreover, it does not state explicitly how costs such as maintenance, labor, etc. were calculated. Without flow sheets, equipment lists, and operating cost assumptions, SERI must take a lot on faith. Unfortunately, this does not put SERI in a good position to speak with authority about the potential of anaerobic digestion.

2 • A more detailed description of the development of the MSWAD model (and therefore the COWSA model from which it was derived) with reference to material costing modules, etc., may increase the confidence of the projections.

3 • If the MSWAD model program is designed to be operated on IBM and compatible micro-computers, use of the program at SERI may be beneficial in ascertaining the usefulness and cost sensitivities of proposed research on the overall economics of the anaerobic process. The program should be supplied with a suitable brief instruction set.

4 • It is unclear whether process designs and cost estimates address the dewatering of lower solids processes such as RefCoM and SOLCON. The disposal of process water may also be a factor if this process water is not completely recycled. Additionally, it

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is unclear whether the process water may be completely recycled. Research by Gaddy and Clausen at Univ. of Arkansas and McCarty at Stanford identified the maximum recycle rate for CSTR reactor systems. I believe the maximum recycle rate was ~70% with inhibition of fermentation at higher recycle rates.

5 • The MSWAD model does not address the need or use of sewage sludge for nutrient supplementation or co-digestion. This was an important topic for studies at the ETU. In our own studies at SERI, we determined that true RDF, in which the wet food and yard waste is removed, is nutrient deficient and will not support effective anaerobic digestion without nutrient supplementation. However, in previous studies with selected MSW which contained all of the food and yard waste, digestion was not nutrient limited. Therefore, the "front-end" system or preprocessing equipment and resulting digester feedstock should be clearly defined to determine if sewage sludge will be necessary for the process.

6 • The operation of MED at high solids levels was a problem. Projections used in the MSWAD model refer to solids levels of 8.7% for the SOLCON system which serves to hamper the use of MED. Furthermore, the use the MED system with the SERI reactor is doubtful as very little free water is available. Estimates from the MSWAD model indicate that direct use of medium BTU gas is the most cost effective of the options examined to date. Obviously this could change, and for low solids systems in regions in where upgraded biogas has a significant market MED may be a strong advantage.

7 • Several sensitivities conducted with the MSWAD model still do not make sense to me, however this does not indicate that they are incorrect. Additionally details concerning the model program in the following areas may be appropriate to enlighten people: levelized gas costs versus solids concentrations, solids retention time, and reactor heating costs.

8 • The recommendations presented in section 4.0 are significant, and substantiated by the information derived from the MSWAD model.

9 • Interesting economic comparison of RefCoM, SOLCON, and SERI High Solids, all at vastly different stages of development and capacities ranging from SERI HS of 2lb/day to 40,000 lb/day for RefCoM. The author's recommendation that SERI HS be scaled up is appropriate based on the favorable economics at the bench scale. A scale of 50-100 lb/day would be in order.

10 • It is not surprising that the SERI HS does not model perfectly as a continuous stirred tank reactor (CSTR.) The CSTR assumes perfect mixing and continuous feed–pretty far removed from the SERI HS operation with intermittent feed and stirring of less that 1 rpm.

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11 • The Argonne Uniform Cost Methodology is widely used in the biofuels arena for costing purposes. How different is Hunter's costing methodology from ANL's, and is there a significant difference in the final outcome?

12 • For the ongoing National Energy Strategy (NES) exercise, digestion of MSW for MBtu gas was projecteu to be ready for the market place by 1995 based on certain assumptions such as a tipping fee of $40/ton. In order to reach this goal, it was assumed that the HS system was demonstrated by 1995. This seems unrealistic considering the state-of-the-technology and all of the uncertainties associated with scale-up. Perusing the Hunter report reinforces this skepticism.

13 • The method of financing and debt/equity reaction has a significant influence on economic analyses; it is not given much treatment in the subject report.

14 • Tipping fee is actually the driver concerning economics for the three processes examined accounting for 79% of the income for SERI HS at $40/ton tipping fee. Recyclable sales is only 7%, indicating that source separation may not impact the MSW situation too dramatically. Recycling by source separation would decrease expendituresfor process equipment, or preprocessing equipment, to be more exact, but would eliminate income from sale. The trade-offs would make for an interesting economic analysis.

15 • Composting of digester residues is treated realistically. As pointed out, marketing hundreds of pounds of compost per day can be a major problem.

16 • Methanol from MSW digester gas is not the most attractive fuel option unless the market for MBtu, SNG, and electric power is exhausted. It is doubtful that the economics for methanol at the scales considered for MSW could ever compete with natural gas as a feedstock or even biomass.

Sincerely,

Barbara Goodman
by K.C.

Barbara Goodman
Senior Project Coordinator
Anaerobic Digestion

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RESPONSE TO SPECIFIC COMMENTS

Note: The responses are numbered according to the numbering system for the comments.

1. Lack of Documentation

This is a systems model, designed to operate over a wide and continuous range of sizes and with many subsystem options to choose from (see Section 2.0 "Model Development," especially Figure 1 on p 3). The general approach has been to develop costs from industry data and to fit cost curves to three, four or more datapoints (see, for example, the development of labor cost in 1988 Annual Report, pp 34, 36 and 37). It was felt that this was the only way to provide systems analysis over the range of parameters requested within the budget provided. This being said, I can only concur that a detailed equipment list would enhance the authority of the report. I also want to apologize if the organization of the report makes it difficult to locate desired information. Hopefully, the listing supplied below will be of some help. Note also that mass and energy balance flow sheets for individual cases can easily be provided, but including flow sheets to cover every case reviewed in the report would have substantially increased the size of the report and made it even less appetizing. The following is a listing of existing documentation:

a) Mass Balance

Rationale for mass balance: pp 6-17
Assumptions (process, operating parameters, economic):

Table 1, p 5;
Table 3, p 24;
pp A-1 through A-4.

Mass balance table: pp 21 and 22; Table 2, p 22
Mass balance flow sheet: pp A-5 through A-7

b) Energy Balance

Rationale for energy balance: pp 17-18
Energy balance table: Table 2, p 22
Energy balance flow sheet: Figure 28, p 65

c) Equipment List

Discussed in general terms on pp 6-17
Listed by group on p A-10.
See also following item

d) Costs

Discussed in general terms on pp 17-18
Listed on p A-10.

Excerpts from 1988 Annual Report:

Rationale for cost calculations: p 29, pp 32 through 45
Cost equations: pp A-49 through A-64

Excerpts from 1986 Equipment Cost Handbook:

Data points and cost curves for some conversion and gas cleanup costs: pp 4-7, 29-31, 45-52

2. Evolution from COWSA to MSWAD

The changes made to COWSA to yield MSWAD are listed on p 4. They include deletions, additions, and internal reorganization and consolidation, but the basic logic as described in the 1988 Annual Report was largely unaffected (see Excerpts from 1988 Annual Report).

3. Software Availability

MSWAD is IBM-compatible. The COWSA instruction manual previously delivered to SERI is usable with minor modifications. A copy of the COWSA manual and a disk with MSWAD are included.

4. Dewatering and Liquid Recycle

The costs of dewatering (capital, labor, maintenance, energy) are included, as can be ascertained from p A-10. The cost of process water disposal is not included because the amount of such waste water is a) small (11,200 gallons per day for a 500-tpd MSW base case, see mass balance flowsheet, p A-6), b) extremely variable depending on feedstock moisture content and biodegradability, percent conversion as defined by operating parameters, and extent of dewatering. The wastewater can be made to disappear entirely as a distinct stream by making minute changes to the feedstock composition, or by dewatering to only 41 percent TS instead of 50 percent TS.

If wastewater is in fact generated, one option is to simply evaporate it as proposed by DRANCO; the amount of heat necessary to do so is equivalent to four percent of the methane stream. If power is cogenerated, the amount of waste heat available is roughly 20 times larger than the amount of heat needed for evaporation. Should the residue be burned, the wastewater can be simply used as ash quenching water.

The statement about recycling limitations is intriguing, and I have never heard of such limitations. Dr. P.L. McCarty thinks that it refers to experiments they did with heat treatment of lignocellulosics whereby toxic compounds are found. Obviously, such would limit the rate of recycle, but does not apply here. More specific information on this topic would be welcome.

5. Nutrient Supplementation

Sewage sludge addition is not necessary because the front end is geared towards producing a refuse-derived feed (which includes food and yard waste), not a refuse-derived fuel. The module design is based on research by J.T. Pfeffer and J.J. Geselbracht (see p 6 of the Excerpts from the 1988 Annual Report). A good summary of their work is provided in: Isaacson, H.R., J.T. Pfeffer, P. Mooij and J.J. Geselbracht, RefCoM--Technical Status, Economics and Market, presented at: Energy from Biomass and Wastes XI, Orlando, Florida, March 16-20, 1987 (Institute of Gas Technology).

6. MED and Solids Content

The comment is well taken: MED requires the presence of a sufficiently dilute liquid to serve as the carrier for CO_2 out of the reactor into the CO_2 stripper. In the case of SOLCON operated with an influent having less than 6.5 percent TS, this liquid can be the "clear" phase at a certain depth in the reactor. In the case of the high solids SEBAC process under development at the University of Florida, a very low solids (< two percent TS) leachate is available for the same purpose.

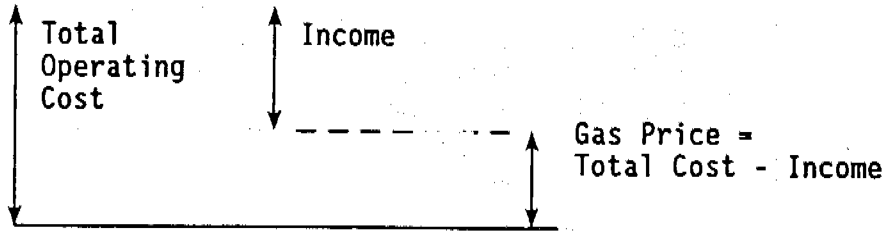
Getting back to the comment though, note that in this report (see p 37, Figure 14 on p 39, and p 4), MED is only applied to the RefCoM base case, having an effluent TS content of eight percent TS.

MED and conventional gas processing were compared for the RefCoM base case. The systems analysis approach of varying one parameter (gas processing) at a time was followed here too. Consequently, the base case slurry solids concentration of eight percent TS was used, which may not be technically compatible with MED. However, the resulting cost comparison between conventional CO_2 removal and MED illustrated in Figure 14, is still valid and will remain largely unchanged if lower solids concentrations are considered. Note also item 8 of the Conclusions of the Summary (page iv) where it is stated that MED has the potential to considerably narrow the cost gap between medium and high Btu gas production.

7. Difficulties with Sensitivity Analyses

7.1 Cost Versus Solids (TS) Content

This relationship is illustrated in Figure 5, p 27 (RefCoM base case) and in Figure 26, p 61 (three processes). The following diagram may clarify Figure 5:



First, the total cost of operation is calculated and divided by the gas production to express it in $/MMBtu of SNG produced. Then, the income derived from tipping fees and aluminum recovery is calculated and also expressed in $/MMBtu. This income is subtracted from the total calculated earlier. The result is the price of gas that must be charged to the customer to break even. The curve in Figure 5 links all these gas prices.

In general, a very dilute process, say a completely mixed reactor operating at three percent TS, requires large reactor volumes to handle a given stream of MSW in a given retention time. This is so because the digester contains mostly water. In the example at three percent TS, for every three tons of dry matter, room must be found in the reactor for 97 tons of water. Large reactors are expensive but costs can be reduced by switching to a more concentrated process because such will reduce the size of the reactor needed and thus the cost. As solids (TS) contents are increased, costs decline, which explains the downward trend of the curves in Figures 5 and 26.

However, reactors are not the only costly items in an MSW biogasification facility. The other costs (preprocessing, postprocessing, gas treatment, etc.) are largely independent of reactor size and will remain constant. This is one of the reasons that as higher and higher TS contents are considered, the decline in the cost curve levels off because the cost of digesters becomes an unimportant part of the total cost.

At high dry matter contents such as experienced in the SERI high solids reactor, another limiting factor enters into play: there is a limit to the amount of dry matter that can be crammed into every unit volume of the reactor. An example may clarify this: let us assume the limit for substrate X is 200 grams dry matter (TS) per liter. Visualize a liter of reactor volume filled to this limit with substrate X. If the reactor is filled with water and substrate X (flooded), the solids content is 20 percent TS (200 g dry matter ÷ approximately 1,000 g aqueous mix per liter). Now consider another one-liter reactor filled with the same material X, but at a solids content of 30 percent TS. It too can only contain 200 grams of dry matter because the solids content of the substrate has (in first approximation) no impact on the dry matter density. In other words, the total reactor volume required to process a given feed flow is the same whether the process is operated at 20 percent TS or at 30 percent TS. This explains the kink in the curve for the high solids reactor when this limiting TS value is reached (see Figure 26). Note that the real relationship between dryness and dry density is more complicated, this is only an approximation.

Once last comment about Figure 26: the curve for RefCoM lies higher than the other curves, reflecting the higher cost due to the energy expense of continuous mixing.

7.2 Cost Versus Retention Time

The longer the retention time, the more feedstock is converted to gas, because the feedstock is exposed to the conversion reaction for a longer duration. So, long retention times maximize gas production. Also, the longer the retention time, the larger the reactors need to be. These are the two main variables impacting cost as retention time is increased. Note that neither of these two relationships is linear with respect to time.

On Table B1, selected datapoints for Figure 3 are listed. The third column contains the total cost (operation, maintenance, and debt service) to operate the plant. Column four contains the credits from tipping fees charged to process MSW, and to a much smaller extent from the sale of recovered aluminum. These credits are subtracted from the costs to yield column 5 which shows remaining cost to be covered by gas sales. By dividing the numbers in this column by the gas production in column 2, the breakeven price of gas is obtained in the last column.

Notice that at very short retention times (six days for example), the total cost is high, because little biogasification occurs, leaving large amounts of residue to be landfilled, which is expensive. Since energy production (SNG) is low, the per-unit price is even more negatively impacted. As longer retention times are considered, total cost (column 3) declines to a minimum around 14 days. However, the per-unit cost ($/MMBtu) continues to decline beyond 14 days SRT, because the gas production is still increasing. The minimum per-unit cost is reached around 23 days. At even longer retention times, the total costs creep upward as larger and larger digesters have to be built while gas production levels off.

7.3 Reactor Heat Balance

A thermal balance for the base case is provided in Table B2. In the first column from the left, the heat balance for the average ambient temperature (55°F) is provided, in the second column for the design temperature (0°F). The digesters are operated at 140°F and are covered with two inches of spray-on polyurethane insulation. In the next two columns, these heat flows are then expressed as a percentage of the energy content of the gross biogas production, which happens to be 100 MMBtu/hr (see also Table 2, p 22, where biogas is referred to as medium Btu gas). The elements of the heat balance are now defined:

Table B1: Explanation of Gas Price Versus Retention Time

(1) SRT (days)	(2) SNG Produced (MMBtu/d)	(3) Total Cost ($/day)	(4) Total Credits ($/day)	(5) = (3) - (4) To be Covered by Gas Sales ($/day)	(5) ÷ (2) Gas Price ($/MMBtu)
6	1,133	28,330	18,530	9,800	8.65
10	1,429	27,535	18,530	9,005	6.30
14	1,610	27,307	18,530	8,777	5.45
18	1,731	27,386	18,530	8,856	5.12
22	1,818	27,641	18,530	9,111	5.01
26	1,884	28,008	18,530	9,478	5.03
30	1,935	28,425	18,530	9,895	5.11
40	2,025	29,704	18,530	11,174	5.52
50	2,083	31,145	18,530	12,615	6.06

a) <u>Gains</u>

Feed enthalpy: Sensible heat of the feedstock, referenced to 0°F.

Mixing heat input: It is assumed that 85 percent of the power consumed by the mixers is converted to friction heat inside the reactor (the remaining 15 percent is dissipated by the motor and drive). There are 15 digesters in the base case, each using 257 Kw for mixing, which translates to 11.20 MMBtu/hr.

Metabolic heat input: Unknown at this point but may be as much as 10 MMBtu/hr.

b) <u>Losses</u>

Wet biogas enthalpy: Sensible heat of biogas saturated with moisture, referenced to 0°F.

Filtercake enthalpy: Sensible heat of dewatered filtercake, referenced to 0°F.

Excess filtrate enthalpy: Sensible heat of the excess filtrate (the part that is not recycled but released from the facility), referenced to 0°F.

Conduction heat loss: Through walls, roof and floor of digester; takes into account resistance coefficient for the interior water film, the steel, the insulation, and the exterior air film. At 55°F outside, the conductive heat loss is 6.39 Btu/sq ft/hr. Individual digesters are 49.7 ft in diameter by 49.7 ft tall.

Recycle heat loss: It is assumed that during dewatering and recycle of the filtrate, the temperature of the filtrate decreases by 10°F.

Evaporative heat loss: The latent heat loss when water evaporates from the digesting slurry into the biogas; it is assumed that the biogas is saturated with moisture.

Note: a radiative heat balance was not made because a) it was calculated that losses would generally be very small (less than one percent of biogas energy content), b) calculating it would have required a half dozen additional specific inputs, thereby unduly complicating the model.

As can be seen from Table B2, mixing energy input overwhelms the heat balance for this RefCoM base case. The calculated overall heat addition is negative, in other words, these reactors may have to be cooled. This may seem outrageous to anyone with experience in anaerobic digestion, but the fact is that nobody has ever biogasified:

o a very dry feedstock,

Table B2: Thermal Balance Around Conversion Stage of RefCoM Base Case

	Average MMBtu/Hr	Design MMBtu/Hr	Percent of Gross Biogas Production Avg.	Percent of Gross Biogas Production Design
Gains:				
Feed enthalpy	1.14	0.00	1.1%	0.0%
Mixing heat input	11.20	11.20	11.2%	11.2%
Metabolic heat input	Unknown	Unknown	N A	N A
Total Gains	12.34	11.20	12.3%	11.2%
Losses:				
Wet biogas enthalpy	0.73	0.73	0.7%	0.7%
Filtercake enthalpy	1.57	1.57	1.6%	1.6%
Excess filtrate enthalpy	0.55	0.55	0.6%	0.6%
Conduction heat loss	0.93	1.54	0.9%	1.5%
Recycle heat loss	1.16	1.16	1.2%	1.2%
Evaporative heat loss	1.47	1.47	1.5%	1.5%
Total Losses	6.41	7.02	6.4%	7.0%
Net heat addition required (= losses minus gains)	-5.93	-4.18	-5.9%	-4.2%

- without any external water addition, but instead internal recycle of liquid,

- in large full-scale digesters with low surface area to volume ratios.

The actual RefCoM experience (Pompano Beach 1978-1985) is not very instructive in this regard in part because heat balances did not warrant a priority on the project. No liquid was recycled but all dilution needs were satisfied by adding cold liquid. Many start-ups occurred (the heat balance presented here presumes steady state).

Note also on Table B2 that metabolic heat production can be as high as 10 MMBtu/hr. So even without any mixing, the metabolic heat input will usually suffice to make the system thermally self-sufficient. Even if there were no metabolic or mixing heat input, the heat addition would only amount to five percent of the energy content of the biogas.

8. Recommendations

No comment.

9. Scale Differences

No comment, except that SERI HS could be safely scaled up to a design capacity of 200 pounds/day in my opinion.

10. SERI HS Modeling Accuracy

Presumably, SERI HS is fed once a day (every 24 hours); the mixer rotates at one rpm. Consequently, over 1,400 revolutions are achieved between feedings and the feed can be assumed to be completely mixed in.

Intermittent feeding could tend to increase average conversion over continuous operation since the mean SRT is increased by removing short circuiting and ensuring a minimum SRT of 24 hours. This is illustrated by the outcome of the simulation attempt (see Table 5, page 56): all performance parameters can be approximated except conversion percent and its corollary, methane yield.

11. Argonne Uniform Cost Methodology

Being unfamiliar with the Argonne Uniform Cost Methodology, I cannot answer this question. For a description of cost methodology, see response to Comment 1 and appendices.

12. Development of MSW Biogasification

On the other hand, the DRANCO system has been in operation at a scale of two tpd in Belgium since 1984, with a larger demonstration facility planned. Negotiations are ongoing with various sponsors to build a SEBAC demonstration plant for operation by 1993.

13. Financing

The method of financing assumed here was municipal ownership with 100-percent debt financing (General Obligation bonds) at 8.1 percent current dollar return. For an analysis of different ownership and financing options, see Isaacson et. al., 1987.

14. Tipping Fee and Source Separation

The tipping fee represents 79 percent of the income for a putative SERI HS facility charging a $40/ton tipping fee. This economic assumption results in a gas price of $1.48/MMBtu, well under market value in many urban areas. A more realistic approach would be to sell the gas at market value, then calculate the breakeven tipping fee. For example, in January 1990, the average industrial gas rate in the U.S. was $3.47/MMBtu; using this number with the SERI HS process results in a tipping fee of $30.5 dollars/ton, responsible for 60 percent of income instead of 79 percent.

Recycling by source separation will reduce the maintenance cost of preprocessing, but probably not its capital cost. This is so because source separation by itself cannot ensure a sufficient feed quality, consequently, one should assume that the same front-end hardware will be required.

Note that income from recyclables was conservatively approached in the base case since only income from the sale of recovered aluminum is considered, at $800/ton (40 cents/pound). The aluminum content of MSW was set at 0.45 percent by mass, a level often observed after a recycling program has been established.

15. Marketing of Compost

No comment except that I wrote that marketing hundreds of tons of compost per day is a problem, not hundreds of pounds.

16. Methanol from Biogas

It is not clear to me why methanol production from MSW-derived biogas could never compete with methanol production from natural gas or biomass. No reasons are given to back up this statement. The present study identifies methanol costs as low as seven cents per gallon if the tipping fee is a competitive $40/ton MSW. I am not aware of costs of methanol production from natural gas or biomass even coming close to this. To be sure, gas or electricity sales are a more straightforward way to market the energy product, but this may change if an urban methanol market develops in the wake of clean air regulations.

METHANE FROM COMMUNITY-DERIVED WASTE SYSTEMS
A&E SUPPORT

ANNUAL REPORT

1988

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TASK I

UPDATE AND DESCRIPTION OF COMMUNITY WASTE MODEL

SECTION 2.0

DESCRIPTION OF MODULES

This chapter describes in some detail the major inputs, calculations, and outputs of the process modules. Actual printouts of the modules in the spreadsheet are shown throughout this section. The basecase assumed for these printouts is summarized on Table 0. For this illustration, the levelized tipping fee version of the model was chosen; gas cost is an input and tipping fee is the output. The equations for each of the modules are included as Appendix A.

2.1 PRIMARY INPUT SECTION

The model user enters values which describe the overall plant configuration in the Input Section. These inputs include sewage flow rate (millions of gallons per day); MSW quality, quantity, and tipping fee or gas price, depending on which version of COWSA was chosen; selection of solid residue processing option; choice of low, average, or high cost estimate for front-end and solid residue processing; and selection of base year for levelized cost. The user also enters values for conversion variables such as solids retention time (SRT), ratio of solids retention time to hydraulic retention time (SRT/HRT), digester temperature, and percentage methane in the biogas produced. These inputs are used in the process modules, and are shown in the computer printout as Figure 6.

2.2 REFUSE SEPARATION MODULE

Kereomel Environmental Systems Analysts, in Champaign, IL, developed a computer model of refuse biogasification which includes a detailed mass balance description of a refuse refining system optimized for anaerobic digestion. From two case studies obtained from Kereomel, equations were derived using ratios of the relative quantities and qualities of the material at specific operating conditions for various stages in the system. A sufficiently accurate mass balance of the front-end system is thus obtained.

The Refuse Separation Module, as shown in Figure 7, accepts MSW as the input, which passes through a shredder, magnets for removal of ferrous metals, disc screens for size classification, an air stoner and air knife for separation of light and heavy materials, aluminum recovery, and plastic/paper separator. The output from the module shows the quantities and qualities of recovered ferrous metals, aluminum, and plastics. The refuse derived fuel (RDF) which results from the separation process is described in some detail, including moisture content, biodegradable and non-biodegradable organics, and inorganics. This constitutes the feed to the digesters. The remainder of the material, largely mineral and unrecyclable residue, can be landfilled directly. The information about the RDF stream is then automatically transferred to the Feed Input Module.

It must be noted that the refuse separation module can be bypassed, for a case where RDF is immediately available. In this case, a negligible MSW inflow can be entered in the primary input section, and the actual RDF amount and quality can be entered directly in the feed input section, as a secondary feedstock (see Section 2.4: Feed Input Module). This should only be done with COWSAGC1, the

The experimental results cast doubt on the feasibility of this technology because it has proved impossible so far to achieve a significant conversion of organics to acid.

2.12 PROCESS ENERGY MODULE

All of the thermal and electric requirements of the total process are calculated in this module, as well as the percentage of gross energy production (in the form of biogas) for each item considered. Based on average and design weather conditions, thermal requirements for the low moisture and high moisture conversion processes are calculated, including influent (feed) heating prior to entry into the digesters, digester heating to maintain the desired temperature level, and thermal drying of filtercake when one of the boiler options is selected. The conductive heat losses through the walls, roof and bottom of all tanks are calculated; all heating needs are calculated for average and design (extreme) conditions. Reactor temperature, average and design ambient temperatures and thickness of sprayed-on polyurethane insulation are inputs. Radiative heat loss was found to be negligible. Evaporative heat loss was not considered although it amounts to around 1% of gross energy (biogas) produced at thermophilic temperature and 0.5% at mesophilic temperature; it may be recovered when the moisture is condensed out during compression and gas cleanup. Piping heat losses are accounted for and are calculated as 15 percent of the tank conductive heat losses. The heating needs estimate is conservative because metabolic heat production (which can amount to several percent of gross energy production) was not included, nor was mixing friction heat in the CSTR option.

It is important to keep in mind that with a dry feed such as MSW, process heating needs are not a concern. In a properly designed system, these process heating needs can be kept down to around 5% of the energy contained in the gross biogas production, irrespective of digester temperature or climate. In fact, after accounting for metabolic and mixing heat, there may not be any need for process heat input. Electricity requirements are determined for all of the process modules including front end power, sewage treatment power, reactor mixing (for CSTR only), pumping and dewatering in the conversion system, mechanical drying, fans and pumps for the boiler plant, and gas compression for cleanup. Total heating requirements in Btu/hr, total electricity in kwh/day, and total percentage of gross biogas production to meet plant energy needs are the results of this module, which is shown in Figures 19(a) and 19(b).

2.13 COST MODULE

The two sections of the Cost Module are the unit cost/credit assumptions and the cost/credit tables, which are both broken down by process module. The assumptions section allows input by the user for unit costs such as value of scrap aluminum, compost, and electricity, and cost of land, electricity, boiler fuel, and landfilling. A printout of this section is included as Figure 20. The cost table contains equations to calculate total costs for the components of the process modules. (See Figure 21.) Capital costs are considered long-term or short-term. Annual costs ($/yr) include operation and maintenance (O&M), and fuel/power. Sewage treatment cost curves were adapted from the Black & Veatch model. Component costs for the high moisture and low moisture conversion and gas cleanup processes were taken from the "Equipment Costs Handbook for Biomass

FIGURE 20

UNIT COST/CREDIT ASSUMPTIONS

1. REFUSE SEPARATION

Shredder	NA /ton
Magnet	NA /ton
Disk Screen	NA /ton
Plastic/Paper Separator	NA /ton
Conveyor	NA /ton
Value of Scrap Iron	$25 /ton
Value of Scrap Aluminum	$500 /ton

2. SEWAGE TREATMENT

Sewage Treatment Credit	$200 /MG sewage
Primary Treatment	see Black & Veatch model cost curves
Hyacinth Ponds	see Black & Veatch model cost curves
Land Cost	$10,000 /acre

3. CONVERSION

A. HIGH MOISTURE SIDE

Digesters	see Equipment Cost Handbook (ECH),section 2
Heating	see ECH, section 3
Dewatering	see ECH, section 7
Insulation	see ECH, section 4

B. LOW MOISTURE SIDE

Digesters	see ECH, section 2
Heating	see ECH, section 3
Dewatering	see ECH, section 7
Mixing	$1.90 /cu ft
Insulation	see ECH, section 4

4. GAS CLEANUP/COMPRESSION

Cleanup	see ECSA documentation
Compression	see ECSA documentation
Value of CO2 recovery	NA

5. SOLID RESIDUE PROCESSING

Landfilling	$20 /ton
Boiler	see model documentation
Turbine	see model documentation
Value of compost	$4 /ton

6. LIQUID RESIDUE PROCESSING

Wastewater treatment	$200 /MG wastewater

7. OTHER

Cost of electricity	$0.06 /kwh
Cost of fuel	$4.00 /MMBtu
Value of electricity	$0.03 /kwh
Value of SNG	$3.00 /MMBtu
Bond financing cost as % of capital	40 %

FIGURE 21

COST TABLE (JUNE 1987 $)

	LONG-TERM CAPITAL ($)	SHORT-TERM CAPITAL ($)	O&M ($/YR)	LABOR ($/YR)	FUEL/POWER ($/YR)	CAPITAL COSTS INCLUDING FINANCING: LONG-TERM CAPITAL ($)	CAPITAL COSTS INCLUDING FINANCING: SHORT-TERM CAPITAL ($)
1. REFUSE SEPARATION	0	9,813,000	205,272	459,760	0	0	12,364,380
2. SEWAGE TREATMENT							
Primary Treatment	4,308,120	1,748,780	324,638		0		
Hyacinth Ponds		2,497,315	72,397		0		
Total	4,308,120	4,246,095	397,035	0	0	5,428,231	5,350,079
3. CONVERSION							
A. HIGH MOISTURE SIDE							
Digesters	582,506		168,761				
Heating	0		0		0		
Dewatering		176,235			0		
Insulation	41,199						
Miscellaneous					0		
Total	623,705	176,235	168,761	0	0	785,868	222,056
B. LOW MOISTURE SIDE							
Digesters	5,208,729		134,242	284,470			
Heating	0		0		0		
Dewatering		2,348,180			0		
Mixing	2,520,332		68,049		0		
Insulation	505,121						
Wet Oxidation/Misc.	0		0		0		
Total	8,234,182	2,348,180	202,291	284,470	0	10,375,069	2,958,707
4. GAS CLEANUP/COMPRESSION							
Gas Cleanup	3,706,149		73,786	102,800	0		
Compression							
Total	3,706,149	0	73,786	102,800	0	4,669,748	0
5. SOLID RESIDUE PROCESSING							
Landfilling			1,174,254				
Boiler	0		0	0	0		
Boiler and Turbine	24,003,492		491,539	1,184,720			
Gasifier	0		0				
Recip engine/generator	0		0				
Delumper/tubgrinder		0	0		0		
Trommel		0	0		0		
Windrow composting		0					
Total	24,003,492	0	1,665,792	1,184,720	0	30,244,400	0
6. LIQUID RESIDUE PROCESSING							
Wastewater Treatment			10,574			0	0
7. GRAND TOTAL	40,875,648	16,583,509	2,723,512	2,031,750	0	51,503,317	20,895,221

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and Waste Systems" compiled by RS&H for GRI. Costs for the refuse separation process (front-end) and residue processing options of boiler plant and boiler/turbine plant (tail-end) were developed in detail for this model. The credit table calculates the annual revenue streams for byproducts and waste disposal. (See Figure 22.) Byproducts are scrap iron, scrap aluminum, compost, and excess electricity generated in the cogeneration option. If electricity requirements by the entire system are not met by an on-site power plant, the cost of electricity for each process module is calculated in the cost table. Waste disposal credits are taken for MSW (as a function of tipping fee) and sewage.

The cost analysis techniques for RDF preparation and solid residue disposal which were described in detail in the RS&H 1987 Annual Report (Legrand, 1988) were evaluated in early 1988. This approach, which uses data from operational and near-operational mass-burn and RDF facilities and applies a statistical treatment, was determined to be inaccurate for O&M and labor costs due to the limited number of observations and a high standard deviation. The capital costs remain the same as described in the 1987 report, with a range of low, average and high values for the boiler in the incineration option, and a cost curve as a function of tons per day of boiler fuel for the boiler/turbine in the incinerator with power generation option.

The O&M and labor costs were revised according to estimates for four plant sizes: 250, 500, 1,000 and 2,000 tpd MSW. Costs were compiled for these sizes and curves were fitted to the points. The O&M costs for the boiler only and the boiler/turbine system consist of two components: spare parts (1.4 percent of capital cost) and consumables, such as lubricants, whose costs are dependent on the quality of material being processed. The labor curves were generated by estimating the total number of plant personnel required for an entire waste-to-energy facility for the various plant sizes and prorating the labor dollars to the capital cost for the four process modules: refuse separation, conversion, gas cleanup, and residue processing. These labor costs are listed for a 1,000 tpd facility in Figures 23(a) and 23(b).

2.14 LEVELIZED COST MODULE

The Levelized Cost Module was developed according to the equations supplied by Decision Focus, Inc., (Clark, 1982). This module receives as inputs the costs and credits of the process modules from the cost/credit tables. The user inputs financial parameters such as cost escalation rates and project funding sources. The costs and credits of each process module for long-term capital, short-term capital, O&M, fuel, land rent, byproduct credit, and waste disposal credit are then levelized to produce either a net total levelized tipping fee or a net total levelized SNG cost, depending on the version of the model being used. The resulting levelized costs include working capital and process development allowances, and represent the price to be charged for the product or service such that the plant will break even. When the value of the SNG is known, the breakeven tipping fee is calculated in terms of dollars per ton; when the tipping fee is an input, the breakeven SNG price is expressed as dollars per MMBtu of SNG. The computer printouts for the Levelized Cost Module are included as Figures 24 (a), (b), and (c).

FIGURE 22

CREDIT TABLE ($/YR)

1. BYPRODUCT CREDIT	
Scrap iron	220,750
Scrap aluminum	719,789
Compost	17,638
CO2 recovery	
Excess electricity generated	379,305
SNG sales	2,724,888
Total	4,062,370
2. WASTE DISPOSAL CREDIT	
Sewage disposal	1,460,000
Total	1,460,000

FIGURE 23(a)

Plant Operating Manpower and Cost

POSITION	PERSONNEL	SHIFTS	TOTAL PERSONNEL	UNIT RATE (1000$/YR)	TOTAL ANNUAL COST (1000$)
PLANT CAPACITY	1000 TPD				
GEN.MGR.	1	1	1	60	60
SECRETARY/CLERK	2	1	2	15	30
ENGINEER	1	1	1	50	50
OPER. SUPT.	1	1	1	48	48
MAINT.SUPT.	1	1	1	48	48
SHIFT SUPV	1	4	4	46	184
SUB TOTAL			10		420
1. REFUSE SEPARATION					
SCALE	1	1	1	35	35
F.E.LOADER	1	1	1	43	43
TIP FLOOR	1	1	1	40	40
EQUIP OPER	3	1	3	42	126
PICKERS/LABORERS	3	1	3	35	105
SUB TOTAL			9		349
2. CONVERSION					
OPERATOR	1	4	4	42	168
3. GAS CLEANUP					
OPERATOR	1	4	4	42	168
4. INCINERATION					
CONTROL ROOM OPER	1	4	4	44	176
BOILER OPERATOR	2	4	8	40	320
JR. OPERATORS	3	4	12	35	420
TURBINE OPERATORS	1	4	4	40	160
SUB TOTAL			28		1076
SUB TOTAL OPERATIONS			55		2181
UNIT COST ($/TON)					$5.98

FIGURE 23(b)

Plant Operating Manpower and Cost

PLANT CAPACITY 1000 TPD

POSITION	PERSONNEL	SHIFTS	TOTAL PERSONNEL	UNIT RATE (1000$/YR)	TOTAL ANNUAL COST (1000$)
5. MAINTENANCE					
MACHINIST	2	1	2	46	92
OILERS	1	1	1	37	37
ELECTRICIAN	2	1	2	49	98
INSTRUMENT	2	1	2	45	90
PIPEFITTERS	2	1	2	50	100
HELPERS	8	1	8	34	272
LABORERS	2	1	2	34	68
SUB TOTAL MAINTENANCE			19		757
GRAND TOTAL			74		2938
UNIT COST ($/TON)					$8.05

FIGURE 24(a)

LEVELIZED COST MODULE

INPUT SECTION

to	initial year of plant operation	1990	fd	fraction financed by debt	0.65
BLA	book life A in years (long term equipment)	30	fce	fraction financed by common equity	0.00
BLB	book life B in years (short term equipment)	10	fpe	fraction financed by preferred equity	0.00
CP	construction period in years	2	fnb	fraction financed by nonborrowed funds	0.35
dc	constant-dollar discount rate	0.05	rd	current-dollar return to debt	0.001
inf	inflation rate	0.06	rce	current-dollar return to common equity	0.000
eb	current $ escalation rate of byproduct value	0.06	rpe	current-dollar return to preferred equity	0.000
ew	current $ esc. rate of waste disposal credit	0.06	rnb	current-dollar return to nonborrowed funds	0.113
ec	current $ construction cost escalation rate	0.06	Tax Rate	combined state and federal income tax rate	0
ef	current $ fuel cost escalation rate	0.06	TLA	tax life A in years (long term equipment)	15
el	current $ land cost escalation rate	0.06	TLB	tax life B in years (short term equipment)	5
em	current $ variable O&M cost escalation rate	0.06	ITC	investment tax credit	0.1
	PDA - Refuse Separation	0.20	WCF	working capital fraction	0.125
	PDA - Sewage Treatment	0.20	SF	service factor	0.94
	PDA - High Moisture Conversion	0.10			
	PDA - Low Moisture Conversion	0.30			
	PDA - Gas Cleanup	0.20			
	PDA - Residue Processing	0.15			

FIGURE 24(b)

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A	number of acres of land	231
DC	design capacity (tons MSW/yr)	219,000
L	land cost per acre (in tb dollars)	10,000
tb	base year for cost estimation	1987

INTERMEDIATE CALCULATIONS

a	allowed AFUDC rate	0.092
d	current-dollar discount rate	0.113
fe	fraction financed by equity	0.350
re	current-dollar return to equity	0.113
ra	weighted-average after-tax cost of capital	0.092
rm	current-dollar return to O&M cost	0.050
rf	current-dollar return to fuel cost	0.050
rb	current-dollar return to byproduct value	0.050
rw	current-dollar return to waste disposal credit	0.050
rl	current-dollar return to land cost	0.050
TC	tax credit	0.000
LR	land rent	1130

CCR CALCULATION

PART A		PART B	
A	0.063	A	0.125
B	0.961	B	0.961
C	0.052	C	0.106
D	0.000	D	0.000
E	0.464	E	0.627
CCRA	0.052	CCRB	0.211

ANNUITY TABLE

Ann(T,d) = annuity factor for T years at rate d

		d						
T		d	dc	rm	rf	rb	rl	rw
		0.113	0.050	0.050	0.050	0.050	0.050	0.050
BLA	30	0.110	0.065	0.065	0.065	0.065	0.065	0.065
BLB	10	0.172	0.130	0.130	0.130	0.130	0.130	0.130

PRESENT WORTH TABLE

PW(T,d) = uniform series present worth for T years at rate d

	d						
T	d	dc	rm	rf	rb	rl	rw
BLA	8.493	15.372	15.372	15.372	15.372	15.372	15.372
BLB	5.016	7.722	7.722	7.722	7.722	7.722	7.722

FIGURE 24(c)

Cost Inputs: (Millions of 1987 $)

	Refuse Separation	Sewage Treatment	Conversion (High H2O)	Conversion (Low H2O)	Gas Cleanup	Residue Processing	Total System
Capital Pt. A	0.000	5.428	0.786	10.375	4.670	30.244	51.503
Capital Pt. B	12.364	5.350	0.222	2.959	0.000	0.000	20.895
O & M - A	0.000	0.200	0.132	0.379	0.177	2.861	3.748
O & M - B	0.205	0.197	0.037	0.108	0.000	0.000	0.548
Fuel A	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Fuel B	0.460	0.000	0.000	0.000	0.000	0.000	0.460
Land Rent	0.261	0.000	0.000	0.000	0.000	0.000	0.261
Byproduct Credit	0.941	0.000	0.018	0.000	0.000	0.379	1.337
Sewage / Gas Credit	0.000	1.460	0.143	2.582	0.000	0.000	4.185
Total	12.350	9.715	1.016	11.239	4.846	32.726	71.893

Levelized Cost: (1987 $/MMBtu)

	Refuse Separation	Sewage Treatment	Conversion (High H2O)	Conversion (Low H2O)	Gas Cleanup	Residue Processing	Total System
Capital Pt. A	0.000	1.377	0.199	2.632	1.185	7.673	13.067
Capital Pt. B	6.637	2.872	0.119	1.588	0.000	0.000	11.217
O & M - A	0.000	0.971	0.639	1.840	0.858	13.898	18.206
O & M - B	0.997	0.957	0.181	0.525	0.000	0.000	2.660
Fuel A	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Fuel B	2.233	0.000	0.000	0.000	0.000	0.000	2.233
Land Rent	1.268	0.000	0.000	0.000	0.000	0.000	1.268
Byproduct Credit	4.569	0.000	0.086	0.000	0.000	1.843	6.497
Sewage / Gas Credit	0.000	7.092	0.694	12.542	0.000	0.000	20.329
Working Capital	0.061	-0.011	0.004	-0.069	0.024	0.228	0.237
PDA	1.327	0.850	0.032	1.266	0.237	1.151	4.863
TOTAL	7.956	-0.075	0.394	-4.760	2.303	21.107	26.926

Levelized Cost Summary: (1987 $/MMBtu)

	Refuse Separation	Sewage Treatment	Conversion (High H2O)	Conversion (Low H2O)	Gas Cleanup	Residue Processing	Total System
Capital Pt.	6.637	4.249	0.319	4.220	1.185	7.673	24.283
O & M	0.997	1.929	0.820	2.365	0.858	13.898	20.866
Fuel	2.233	0.000	0.000	0.000	0.000	0.000	2.233
Land Rent	1.268	0.000	0.000	0.000	0.000	0.000	1.268
Byproduct Credit	4.569	0.000	0.086	0.000	0.000	1.843	6.497
Sewage / Gas Credit	0.000	7.092	0.694	12.542	0.000	0.000	20.329
Working Capital	0.061	-0.011	0.004	-0.069	0.024	0.228	0.237
PDA	1.327	0.850	0.032	1.266	0.237	1.151	4.863
TOTAL COST	12.525	7.017	1.174	7.782	2.303	22.950	53.752
TOTAL CREDIT	4.569	7.092	0.780	12.542	0.000	1.843	26.826

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2.15 LEVELIZED COST THEORY

The cost analysis of the entire system is based on a levelized cost-of-service price methodology provided by GRI. The cost-of-service price is a price per unit of methane ($/MMBtu) sufficient to generate revenues to meet the following requirements:

1. Amortize debt,
2. Cover operating and maintenance costs and fuel expenditures, and
3. Provide a return on both common and preferred equity.

The levelized cost-of-service price represents a constant dollar per unit price which, if charged for each unit of output over the life of the plant, would yield the same revenue value as would the actual cost-of-service price, discounted to its present value. Thus, the current dollar cost-of-service price is discounted to its present value and levelized over the life of the plant using a constant dollar annuity factor.

The levelized cost for each process is calculated from the total plant investment, variable operating and maintenance cost, fuel expenditures, income taxes, and working capital. The total plant investment is the sum of all construction costs, including site preparation and improvements; plant and process costs; and indirect costs (sales tax on materials, contingency funds, contractor overhead and fees, engineering and design costs, and cost of spare parts). The total plant investment is allocated to the expected plant output using the product of the service factor and net output, resulting in the specific plant investment expressed in dollars per MMBtu per year.

The variable operating and maintenance cost (VOM) has been estimated for the base year and allocated to the expected output. Working capital (WC) has been estimated as a constant fraction of total revenues received each year. Income taxes (where appropriate) have been computed as a function of revenues and are combined at the state and federal tax rate.

The specific plant investment (SPI) is used with the capital charge rate (CCR) to identify the capital charged to each unit of production for long-term and short-term equipment. The unit variable operating and maintenance cost, the unit fuel cost and the unit land rental cost are then converted to their present worth and levelized using the appropriate annuity factor. The levelized unit working capital factor, adjusted by the weighted-average after tax cost of capital and the tax rate, is determined from the unit capital investment, the levelized unit variable operating and maintenance cost, and the levelized unit fuel expenditures. Credits for byproducts, such as scrap aluminum, are accounted for and may be included in the analysis if desired. A process development allowance (PDA) may also be used; the PDA accounts for an increase in the system costs as they move from one level of development to another, and as the definition of the process becomes more detailed.

The level unit cost is calculated as the product of the capital charge rate and the specific plant investment, plus the levelized unit variable factors, operating and maintenance, fuel expenditures, land costs, and working capital. This relationship is expressed symbolically as:

$$LC = (SPI * CCR) + \overline{VOM} + \overline{FUEL} + \overline{WC} + \overline{LAND} - \overline{BPC} + \overline{PDA}$$

where: LC = levelized cost,

SPI = specific plant investment, the plant investment per unit yearly output,

CCR = capital charge rate,

$\overline{VOM}$ = levelized variable operating and maintenance cost per unit output,

$\overline{FUEL}$ = levelized fuel cost per unit output,

$\overline{WC}$ = levelized cost of working capital per unit output,

$\overline{LAND}$ = levelized cost of working capital per unit output,

$\overline{BPC}$ = levelized value of the byproduct credit per unit primary product output, and

$\overline{PDA}$ = process development allowance.

Note that the Capital Charge Rate (CCR) requires the solution of a complex equation, which was split up into five parts in the model.

Each of these elements of the levelized cost of service is computed fr[om] the basic data describing a plant.

$$SPI = \frac{TPI}{SF * DC}$$

$$CCR = \frac{Ann(BL,d_c)}{1-Tax\ Rate}\left(\frac{1}{1+e_c}\right)^{CP/3}\left(\frac{1+e_c}{1+inf}\right)^{t_o-t_b}$$

$$* \left\{ [(1+a)^{CP/3} - itc] \left[\frac{r_a}{d} + \frac{PW(BL,d)}{BL}\left(1-\frac{r_a}{d}\right)\right]\right.$$

$$- Tax\ Rate \left(1 - \frac{itc}{2}\right)$$

$$\left. * \left[\frac{r_a}{d}\left(\frac{1.5 + d\ TL\ \frac{(1-1.5/TL)^{TL}}{1+d}}{d\ TL + 1.5}\right) + \left(\frac{PW(BL,d)}{BL}\right)\left(1-\frac{r_a}{d}\right)\right]\right\}$$

$$\overline{VOM} = Ann(BL,d_c)\left(\frac{1+e_m}{1+inf}\right)^{t_o-t_b} PW(BL,r_m)\ \frac{VOM}{SF * DC}$$

$$\overline{Fuel} = Ann(BL,d_c)\left(\frac{1+e_f}{1+inf}\right)^{(t_o-t_b)*} PW(BL,r_f)\ \frac{FC}{SF * DC}$$

$$\overline{BPC} = Ann(BL,d_c)\left(\frac{1+e_b}{1+inf}\right)^{t_o-t_b} PW(BL,r_b)\ \frac{BPC}{SF * DC}$$

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* The $(t_o - t_b)$ term in the $\overline{\text{Fuel}}$ equation was set equal to 1.0 per GRI request.

$$\overline{WC} = \frac{r_a}{1 - \text{Tax Rate}} \quad WCF * [SPI * CCR + \overline{VOM} + \overline{Fuel} - \overline{BPC}]$$

$$\overline{LAND} = Ann(BL, d_c) \quad \left(\frac{1 + e_l}{1 + inf}\right)^{t_o - t_b - CP} \quad PW(BL, r_l) \left(\frac{A * LR}{SF * DC}\right)$$

where:

a = allowed AFUDC rate,

A = number of acres of land,

Ann(T,d) = annuity factor for T years at rate d,

BL = book life,

BPC = first year byproduct credit (in t_b dollars),

CP = construction period, years,

d = current-dollar discount rate,

d_c = constant-dollar discount rate,

DC = design capacity (MMBtu/yr),

e_b = current-dollar escalation rate of byproduct value,

e_c = current-dollar construction cost escalation rate,

e_f = current-dollar fuel cost escalation rate,

e_l = current-dollar land cost escalation rate,

e_m = current-dollar variable operating and maintenance cost escalation rate,

FC = total first year cost of fuel (in t_b dollars,

f_d = fraction financed by debt,

f_e = fraction financed by equity,

inf = inflation rate,

LTC = investment tax credit,

L = land cost per acre in t_b dollars,

PW(T,d) = uniform series present worth factor for T years at rate d,

EAR17

r_a = weighted-average after-tax cost of capital,

r_d = return to debt,

r_e = return to equity,

SF = service factor,

Tax Rate = combined state and federal income tax rate,

t_b = base year for cost estimation,

TL = tax life, years,

t_o = year of initial operation,

TPI = total plant investment (in t_b dollars),

VOM = total first year variable operating and maintenance cost (in t_b dollars),

WCF = working capital fraction,

and

$$Ann(T,d) = \frac{d(1+d)^T}{(1+d)^T-1}$$

$$PW(T,d) = [Ann(T,D)]^{-1}$$

$$d = [(1 + d_c)(1 + inf) - 1]$$

$$LR = L(d + TC),$$

$$TC = (d_c - r_d f_d) \times \text{Tax Rate},$$

$$r_a = r_e f_e + (1\text{-Tax Rate}) r_d f_d,$$

$$r_m = (d-e_m)/(1+e_m),$$

$$r_f = (d-e_f)/(1+e_f),$$

$$r_b = (d-e_b)/(1+e_b).$$

$$r_l = (d-e_l)/(1+e_l).$$

APPENDIX A

DOCUMENTATION OF EQUATIONS

IN COWSA MODEL

COST & CREDIT TABLE (CT)

1. UNIT COST/CREDIT ASSUMPTIONS

CT.1.1 tipping fee for waste disposal in $/ton = MSW TIPPING FEE IN 1987 $/TON FROM PRIMARY INPUT SECTION

2. ENR COST INDEXES

CT.2.1 Selected year index ratio :

a) If BASE YEAR DOLLARS = 1980 , index ratio = index value ratio for June '80

b) If BASE YEAR DOLLARS = 1981 , index ratio = index value ratio for June '81

ETC.

CT.2.2 index value escalation rate = INFLATION RATE FROM LEVELIZED COST MODULE

CT.2.3 1980 index value ratio = JUNE 1980 INDEX VALUE / JUNE 1987 INDEX VALUE

CT.2.4 1981 index value ratio = JUNE 1981 INDEX VALUE / JUNE 1987 INDEX VALUE

ETC.

CT.2.5 1990 index value ratio = JUNE 1990 INDEX VALUE / JUNE 1987 INDEX VALUE

3. COST TABLE - REFUSE SEPARATION

CT.3.1 TOTAL LONG-TERM CAPITAL COST IN $ = 0

CT.3.2 total short-term capital cost in $:

a) If low cost estimate is chosen, total short-term capital cost in $ = $4773/ton MSW x MSW wet tpd from intermediate summary

b) If avg cost estimate is chosen, total short-term capital cost in $ = $16,355/ton MSW x MSW wet tpd from intermediate summary

c) If high cost estimate is chosen, total short-term capital cost in $ = $27,937/ton MSW x MSW wet tpd from intermediate summary

CT.3.3 total O&M cost in $/yr = 0.014 x total short-term capital cost in $ + ($0.31/ton MSW x MSW WET TPD FROM INPUT SECTION x 365 days/yr)

CT.3.4 total labor cost in $/yr = [143.2 + 0.617 x MSW WET TPD - 0.000149 x (MSW WET TPD)^2] x 1000

CT.3.5 total fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, total fuel/power cost in $/yr = 0

b) Otherwise, total fuel/power cost in $/yr = avg front end power in kwh/day from energy module x cost of electricity in $/kwh x 365 days/yr

CT.3.6 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.3.7 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

4. COST TABLE - SEWAGE TREATMENT

CT.4.1 primary treatment long-term capital cost in $ = [418.6 + 229.9 x mgd sewage - 2.048 x (mgd sewage)^2 + 0.01384 x (mgd sewage)^3] x 1000, where mgd sewage is from sewage treatment module

CT.4.2 primary treatment short-term capital cost in $ = [171.1 + 95.68 x mgd sewage - 1.075 x (mgd sewage)^2 + 0.01176 x (mgd sewage)^3] x 1000, where mgd sewage is from sewage treatment module

CT.4.3 primary treatment O&M cost in $/yr = [50.16 + 15.6 x mgd sewage - 0.1128 x (mgd sewage)^2 + 0.0009498 x (mgd sewage)^3] x 1000

CT.4.4 primary treatment fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, primary treatment fuel/power cost in $/yr = 0

b) Otherwise, primary treatment fuel/power cost in $/yr = [4.42 + 3.687 x mgd sewage - 0.02125 x (mgd sewage)^2 + 0.0001939 x (mgd sewage)^3] x 1000

CT.4.5 hyacinth ponds short-term capital cost in $ = [129.6 + 13.51 x pond acres + 0.007501 x (pond acres)^2 - 9.686E-6 x (pond acres)^3] x 1000, where pond acres is from sewage treatment module

CT.4.6 hyacinth ponds O&M cost in $/yr = [-0.7712 + 0.4535 x pond acres - 0.00003 x (pond acres)^2 - 4.40E-8 x (pond acres)^3] x 1000

CT.4.7 hyacinth ponds fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, hyacinth ponds fuel/power cost in $/yr = 0

b) Otherwise, hyacinth ponds fuel/power cost in $/yr = [-0.3179 + 0.2338 x pond acres - 3.454E-5 x (pond acres)^2 + 9.28E-8 x (pond acres)^3] x 1000

CT.4.8 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.4.9 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

5. COST TABLE - HIGH MOISTURE CONVERSION

CT.5.1 digesters long-term capital cost in $ = [99.29 x (high moisture usable vol per digester in cf from intermediate summary)^0.705] x (high moisture no. of digesters from intermediate summary + 1) / Jan 1985 cost index ratio

CT.5.2 digesters O&M cost in $/yr = [1109 x (high moisture usable vol per digester in cf from intermediate summary)^0.363] x (high moisture no. of digesters from intermediate summary + 1) / Jan 1985 cost index ratio

CT.5.3 heating long-term capital cost in $:

a) If landfill option is chosen, heating long-term capital cost in $ = 20861 x [(design high moisture influent heating + design high moisture digester heating in 1000 Btu/hr from energy module) x 1000]^0.757

b) Otherwise, heating long-term capital cost in $ = 0

CT.5.4 heating O&M cost in $/yr :

a) If landfill option is chosen, heating O&M cost in $/yr = 1814.5 x [(design high moisture influent heating + design high moisture digester heating in 1000 Btu/hr from energy module) x 1000]^0.757

b) Otherwise, heating O&M cost in $/yr = 0

CT.5.5 heating fuel/power cost in $/yr :

a) If landfill option is chosen, heating fuel/power cost in $/yr = (avg high moisture influent heating + avg high moisture digester heating in 1000 Btu/hr from energy module) / 1000 / 0.8 boiler efficiency x 8760 hr/yr x cost of fuel in $/MMBtu from unit cost assumptions

b) Otherwise, heating fuel/power cost in $/yr = 0

CT.5.6 dewatering short-term capital cost in $ = $15,000/dry tpd x high moisture dry tpd to be dewatered from intermediate summary

CT.5.7 dewatering fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, dewatering fuel/power cost in $/yr = 0

b) Otherwise, dewatering fuel/power cost in $/yr = avg high moisture mechanical drying in kwh/day from energy module x 365 days/yr x cost of electricity in $/kwh

CT.5.8 insulation long-term capital cost in $ = high moisture no. of digesters x aboveground surface area per digester in sf from intermediate summary x (1.53 + 0.916 x insulation thickness in inches) $/sf

CT.5.9 miscellaneous fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, miscellaneous fuel/power cost in $/yr = 0

b) Otherwise, misc. fuel/power cost in $/yr = avg high moisture other hydraulic in kwh/day x cost of electricity in $/kwh x 365 days/yr

CT.5.10 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.5.11 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

6. COST TABLE - LOW MOISTURE CONVERSION

CT.6.1 digesters long-term capital cost in $ = [99.29 x (low moisture usable vol per digester in cf from intermediate summary)^0.785] x (low moisture no. of digesters from intermediate summary + 1) / Jan 1985 cost index ratio

CT.6.2 digesters O&M cost in $/yr = 0.014 x digesters long-term capital cost in $
+ ($0.28/ton MSW x MSW WET TPD FROM INPUT SECTION x 365 days/yr)

CT.6.3 digesters labor cost in $/yr = [-11.33 + 0.562 x MSW WET TPD - 0.000115 x (MSW WET TPD)^2] x 1000

CT.6.4 heating long-term capital cost in $:

a) If landfill option is chosen, heating long-term capital cost in $ = 20861 x [(design low moisture influent heating
+ design low moisture digester heating in 1000 Btu/hr from energy module) x 1000]^0.757

b) Otherwise, heating long-term capital cost in $ = 0

CT.6.5 heating O&M cost in $/yr :

a) If landfill option is chosen, heating O&M cost in $/yr = 1814.5 x [(design low moisture influent heating
+ design low moisture digester heating in 1000 Btu/hr from energy module) x 1000]^0.757

b) Otherwise, heating O&M cost in $/yr = 0

CT.6.6 heating fuel/power cost in $/yr :

a) If landfill option is chosen, heating fuel/power cost in $/yr = (avg low moisture influent heating
+ avg low moisture digester heating in 1000 Btu/hr from energy module) / 1000 / 0.8 boiler efficiency
x 8760 hr/yr x cost of fuel in $/MMBtu from unit cost assumptions

b) Otherwise, heating fuel/power cost in $/yr = 0

CT.6.7 dewatering short-term capital cost in $ = $15,000/dry tpd x low moisture dry tpd to be dewatered from intermediate summary

CT.6.8 dewatering fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, dewatering fuel/power cost in $/yr = 0

b) Otherwise, dewatering fuel/power cost in $/yr = avg low moisture mechanical drying in kwh/day from energy module
x 365 days/yr x cost of electricity in $/kwh

CT.6.9 mixing long-term capital cost in $ = MIXING COST ASSUMPTION IN $/CF x low moisture slurry vol in cf from intermediate summary

CT.6.10 mixing O&M cost in $/yr = 0.027/yr x mixing long-term capital cost in $

CT.6.11 mixing fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, mixing fuel/power cost in $/yr = 0

b) Otherwise, mixing fuel/power cost in $/yr = avg entire plant reactor mixing in kwh/day x 365 days/yr x cost of electricity in $/kwh

CT.6.12 insulation long-term capital cost in $ = low moisture no. of digesters x aboveground surface area per digester in sf from intermediate summary x (1.53 + 0.916 x insulation thickness in inches) $/sf

CT.6.13 wet oxidation/misc. long-term capital cost in $:

a) If wet ox. option is chosen, wet ox./misc. long-term capital cost in $ = 101.23 x (reactor effluent in gpd)^0.83740 + 33.67 x (wet ox. ww digester total vol in cf)^ 0.803

where reactor effluent in gpd is from wet oxidation module

b) Otherwise, wet oxidation/misc. long-term capital cost in $ = 0

CT.6.14 wet oxidation O&M cost in $/yr :

a) If wet oxidation option is chosen, wet ox./misc. O&M cost in $/yr = 9.051 x (reactor effluent in gpd)^0.6222

b) Otherwise, wet oxidation/misc. O&M cost in $/yr = 0

CT.6.15 wet oxidation/misc. fuel/power cost in $/yr = misc. fuel/power cost in $/yr + wet oxidation fuel/power cost in $/yr

where misc. fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, miscellaneous fuel/power cost in $/yr = 0

b) Otherwise, misc. fuel/power cost in $/yr = avg low moisture other hydraulic in kwh/day x cost of electricity in $/kwh x 365 days/yr

where wet oxidation fuel/power cost in $/yr :

a) If wet ox. option is chosen, wet ox. fuel/power cost in $/yr = avg wet ox comp & pump from energy module x cost of electricity in $/kwh x 365 days/yr

b) Otherwise, wet ox. fuel/power cost in $/yr = 0

CT.6.16 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.6.17 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

7. COST TABLE - GAS CLEANUP/COMPRESSION

CT.7.1 gas cleanup long-term capital cost in $:

a) If Prism membrane gas cleanup system is chosen, gas cleanup long-term capital cost in $ = (1.0853 - 0.007813 x avg biogas quality in % methane from conversion summary) x total biogas flow in MMscf/day x 10E6 + $200,000

b) If Zeolite process is chosen, gas cleanup long-term capital cost in $ = $1.091/(scf/day) x total biogas flow in MMscf/day from intermediate summary x 10E6

CT.7.2 gas cleanup O&M cost in $/yr :

a) If Prism membrane system is chosen, gas cleanup O&M cost in $/yr = 0.014/yr x gas cleanup long-term capital cost in $ + $0.1/ton MSW x MSW WET TPD FROM INPUT SECTION x 365 days/yr

b) If Zeolite system is chosen, gas cleanup O&M cost in $/yr = 0.035/yr x gas cleanup long-term capital cost in $

CT.7.3 gas cleanup labor cost in $/yr :

a) If Prism membrane system is chosen, gas cleanup labor cost in $/yr = (12.2 + 0.151 x MSW WET TPD FROM INPUT SECTION) x 1000

b) If Zeolite system is chosen, gas cleanup labor cost in $/yr = 0

CT.7.4 gas cleanup fuel/power cost in $/yr :

a) If boiler/turbine option is chosen, gas cleanup fuel/power cost in $/yr = 0

b) Otherwise, for Prism membrane system :

gas cleanup fuel/power cost in $/yr = avg gas cleanup energy consumption in kwh/day from energy module x cost of electricity in $/kwh x 365 days/yr

c) Otherwise, for Zeolite process :

gas cleanup fuel/power cost in $/yr = Zeolite process kw from gas cleanup module x cost of electricity in $/kwh x 8760 hr/yr

CT.7.5 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.7.6 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

8. COST TABLE - SOLID RESIDUE PROCESSING

CT.8.1 landfilling O&M cost in $/yr :

a) If landfill option is chosen, landfilling cost in $/yr = (wet tpd to landfill from refuse separation module + low moisture filtercake (compost) wet tpd from intermediate summary) x LANDFILL COST ASSUMPTION IN $/TON x 365 days/yr

b) If boiler only option is chosen, landfilling cost in $/yr = (wet tpd to landfill from refuse separation module + wet tpd ash to landfill from burning, no power option) x LANDFILL COST ASSUMPTION IN $/TON x 365 days/yr

c) If boiler/turbine option is chosen, landfilling cost in $/yr = (wet tpd to landfill from refuse separation module + wet tpd ash to landfill from burning, power generation module) x LANDFILL COST ASSUMPTION IN $/TON x 365 days/yr

d) If wet oxidation option is chosen, landfilling cost in $/yr = (wet tpd to landfill from refuse separation module + filtercake wet tpd from wet oxidation module) x LANDFILL COST ASSUMPTION IN $/TON x 365 days/yr

CT.8.2 boiler long-term capital cost in $:

a) If boiler only option is chosen,

1) If low cost estimate is chosen, boiler long-term capital cost in $ =
$34,436/wet tpd x total boiler fuel wet tpd from intermediate summary

2) If avg cost estimate is chosen, boiler long-term capital cost in $ =
$58,145/wet tpd x total boiler fuel wet tpd from intermediate summary

3) If high cost estimate is chosen, boiler long-term capital cost in $ =
$81,854/wet tpd x total boiler fuel wet tpd from intermediate summary

b) Otherwise, boiler long-term capital cost in $ = 0

CT.8.3 boiler O&M cost in $/yr :

a) If boiler only option is chosen, boiler O&M cost in $/yr = 0.014/yr x boiler long-term capital cost in $
+ 0.71 x MSW WET TPD FROM INPUT SECTION x 365 days/yr

b) Otherwise, boiler O&M cost in $/yr = 0

CT.8.4 boiler labor cost in $/yr :

a) If boiler only option is chosen, boiler labor cost in $/yr = [-16 + 2.468 x MSW WET TPD - 0.000778 x (MSW TPD)^2] x 1000

b) Otherwise, boiler labor cost in $/yr = 0

CT.8.5 boiler fuel/power cost in $/yr :

a) If boiler only option is chosen, boiler fuel/power cost in $/yr = avg solid residue processing fans & pumps energy consumption
in kwh/day x cost of electricity in $/kwh x 365 days/yr

b) Otherwise, boiler fuel/power cost in $/yr = 0

CT.8.6 boiler/turbine long-term capital cost in $:

a) If boiler/turbine option is chosen, boiler/turbine long-term capital cost in $ = 156140 x (total boiler fuel wet tpd
from intermediate summary)^0.911

b) Otherwise, boiler/turbine long-term capital cost in $ = 0

CT.8.7 boiler/turbine O&M cost in $/yr :

a) If boiler/turbine option is chosen, boiler/turbine O&M cost in $/yr = 0.014/yr x boiler/turbine long-term capital cost in $
+ 0.71 x MSW WET TPD FROM INPUT SECTION x 365 days/yr

b) Otherwise, boiler/turbine O&M cost in $/yr = 0

CT.8.8 boiler/turbine labor cost in $/yr :

a) If boiler/turbine option is chosen, boiler/turbine labor cost in $/yr = [-16 + 2.468 x MSW WET TPD - 0.000778 x (MSW WET TPD)^2]
x 1000

b) Otherwise, boiler/turbine labor cost in $/yr = 0

CT.8.9 gasifier long-term capital cost in $:

a) If gasification option is chosen, gasifier long-term capital cost = 6,030,000 + 46,900 x gasifier fuel dry tpd

b) Otherwise, gasifier long-term capital cost = 0

CT.8.10 gasifier O&M cost in $/yr :

a) If gasification option is chosen, gasifier O&M cost = 34,049 + 1051 x gasifier fuel dry tpd

b) Otherwise, gasifier O&M cost = 0

CT.8.11 reciprocating engine/generator long-term capital cost in $:

a) If gasification option is chosen, recip engine/generator long-term capital cost = 500 x power output of engine in kw

b) Otherwise, recip engine/generator capital cost = 0

CT.8.12 reciprocating engine/generator O&M cost in $/yr :

a) If gasification option is chosen, recip engine/generator O&M cost = 0.01 x power output of engine in kwh/day x 365 days/yr

b) Otherwise, recip engine/generator O&M cost = 0

CT.8.13 delumper/tubgrinder short-term capital cost in $:

a) If composting option is chosen, delumper/tubgrinder capital cost = 1033 x raw compost wet tpd

b) Otherwise, delumper/tubgrinder capital cost = 0

CT.8.14 delumper/tubgrinder O&M cost in $/yr :

a) If composting option is chosen, delumper/tubgrinder O&M cost = 1040 x raw compost wet tpd

b) Otherwise, delumper/tubgrinder O&M cost = 0

CT.8.15 delumper/tubgrinder fuel/power cost in $/yr :

a) If composting option is chosen, delumper/tubgrinder fuel/power cost = 260 x raw compost wet tpd

b) Otherwise, delumper/tubgrinder fuel/power cost = 0

CT.8.16 trommel short-term capital cost in $:

a) If composting option is chosen, trommel capital cost = 1850 x raw compost wet tpd + 0.703 x (raw compost wet tpd)^2

b) Otherwise, trommel capital cost = 0

CT.8.17 trommel O&M cost in $/yr :

a) If composting option is chosen, trommel O&M cost = 52 x raw compost wet tpd

b) Otherwise, trommel O&M cost = 0

CT.8.18 trommel fuel/power cost in $/yr :

a) If composting option is chosen, trommel fuel/power cost = 104 x raw compost wet tpd x cost of electricity in $/kwh

b) Otherwise, trommel fuel/power cost = 0

CT.8.19 windrow composting short-term capital cost in $:

a) If composting option is chosen, windrow composting capital cost = 15 x total floor area for compost in sf

b) Otherwise, windrow composting capital cost = 0

CT.8.20 total long-term capital cost including financing in $ = total long-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

CT.8.21 total short-term capital cost including financing in $ = total short-term capital cost in $ x FRACTION FINANCED BY DEBT x (1 + BOND FINANCING COST AS A % OF CAPITAL / 100)

9. COST TABLE - LIQUID RESIDUE PROCESSING

CT.9.1 wastewater treatment O&M cost in $/yr = WASTEWATER TREATMENT COST IN $/MG WASTEWATER x gpd wastewater / 1,000,000 x 365 days/yr

a) If wet oxidation option is chosen, gpd wastewater = wastewater in gpd from high moisture conversion + (excess filtrate in gpd - dilution water in gpd) from wet oxidation module

b) If wet oxidation option is not chosen, gpd wastewater = wastewater in gpd from high moisture conversion + (excess filtrate in gpd - dilution water in gpd) from low moisture conversion module

10. COST TABLE - GRAND TOTALS

CT.10.1 grand total long-term capital cost in $ = long-term capital costs for refuse separation, sewage treatment, high moisture conversion, low moisture conversion, gas cleanup/compression, solid residue processing, and liquid residue processing

ETC.

11. CREDIT TABLE - BYPRODUCT CREDIT

CT.11.1 scrap iron credit in $/yr = VALUE OF SCRAP IRON IN $/TON x tpd Fe from two magnets from refuse separation module x 365 days/yr

CT.11.2 scrap aluminum credit in $/yr = VALUE OF SCRAP ALUMINUM IN $/TON x tpd Al from refuse separation module x 365 days/yr

CT.11.3 compost credit in $/yr = VALUE OF COMPOST IN $/TON x high moisture filtercake (compost) wet tpd from intermediate summary x 365 days/yr

CT.11.4 credit for excess electricity generated/cost of supplemental power in $/yr :

a) If boiler/turbine option is chosen,

1) If excess power in kwh/day from boiler/turbine module > 0, credit for excess electricity in $/yr = excess power in kwh/day x VALUE OF ELECTRICITY IN $/KWH x 365 days/yr

2) Otherwise, cost of supplemental electricity in $/yr = supplemental power in kwh/day x COST OF ELECTRICITY IN KWH/DAY x 365 days/yr

b) Otherwise, credit/cost = 0

CT.11.5 SNG sales in $/yr = VALUE OF SNG IN $/MMBTU x net production of SNG in TBtu/yr from intermediate summary x 10^{6}

12. CREDIT TABLE - WASTE DISPOSAL CREDIT

CT.12.1 sewage disposal credit in $/yr = SEWAGE TREATMENT CREDIT ASSUMPTION IN $/MG SEWAGE x MGD SEWAGE FROM INPUT SECTION x 365 days/yr

APPENDIX D

EXCERPTS FROM

1986 EQUIPMENT COST HANDBOOK

0.1.11 low moisture VSLR in lb VS/cf-day :

a) If wet oxidation option is chosen, low moisture VSLR in lb VS/cf-day = VSLR in lb VS/cf-day from wet oxidation conversion module

b) Otherwise, low moisture VSLR in lb VS/cf-day = VSLR in lb VS/cf-day from low moisture conversion module

0.1.12 high moisture % of gross methane production = high moisture gross production in TBtu/yr / total gross production in TBtu/yr from intermediate summary x 100

0.1.13 low moisture % of gross methane production = low moisture gross production in TBtu/yr / total gross production in TBtu/yr from intermediate summary x 100

2. LEVELIZED COST OUTPUTS

0.2.1 base year for levelized cost = BASE YEAR DOLLAR INPUT

0.2.2 refuse separation levelized cost in $/ton = selected year index ratio x refuse separation total cost in $/ton from levelized cost summary

0.2.3 sewage treatment levelized cost in $/ton = selected year index ratio x sewage treatment total cost in $/ton from levelized cost summary

0.2.4 high moisture conversion levelized cost in $/ton = selected year index ratio x high moisture conversion total cost in $/ton from levelized cost summary

0.2.5 low moisture conversion levelized cost in $/ton = selected year index ratio x low moisture conversion total cost in $/ton from levelized cost summary

0.2.6 gas cleanup levelized cost in $/ton = selected year index ratio x gas cleanup total cost in $/ton from levelized cost summary

0.2.7 solid residue processing levelized cost in $/ton = selected year index ratio x solid residue processing total cost in $/ton from levelized cost summary

0.2.8 total levelized cost in $/ton = sum of levelized costs of process modules in $/ton

0.2.9 refuse separation fraction of total cost = refuse separation levelized cost in $/ton / total levelized cost in $/ton

ETC.

0.2.9 levelized gas credit in $/ton = - selected year index ratio x (high moisture conversion sewage/gas credit in $/ton + low moisture conversion sewage/gas credit in $/ton from levelized cost summary)

0.2.10 levelized sewage credit in $/ton = - selected year index ratio x sewage treatment sewage/gas credit in $/ton from lev. cost summary

0.2.11 levelized byproduct credit in $/ton = - selected year index ratio x total system byproduct credit in $/ton from levelized cost summar

0.2.12 levelized total credit in $/ton = sum of levelized credits of gas, sewage, and byproducts in $/ton

0.2.13 net total levelized tipping fee in $/ton = total levelized cost in $/ton + levelized total credit in $/ton

THE METHANE FROM BIOMASS AND WASTE PROGRAM

TASK IV: EQUIPMENT COSTS HANDBOOK

FOR

BIOMASS AND WASTE SYSTEMS

VOLUME I

SUMMARY TABLES

AND COST EQUATIONS

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ECH2

2.1 TANKS AND MATERIAL HANDLING

Unit costs of tanks and material handling for Leach-Bed (LB) reactors are shown as a function of usable tank volumes and of material handling rates (loading and unloading). Tank costs include: (1) tank, (2) rock bed, (3) distribution piping support steel, (4) collection piping, (5) distribution piping, (6) washout piping, (7) washout pump, (8) platforms, (9) unloading piping, (10) unloading pumps and (11) loading conveyor.

Tank Shell

- o Usable tank volume is the maximum volume of the tank occupied by biomass feedstock.
- o Tank configuration: circular, containing 4 ft. of sand and rock to provide filtering for the liquid and 5 ft. of freeboard.
- o Types of tank construction considered:
  - Bolted Steel, uninsulated*
  - Prestressed Poured Concrete
  - Hypalon-lined Earthen Wall

Loading and Unloading

- o Rates of material loading and unloading considered: 12, 24 and 48 hours.
- o One solids movement system per 6 tanks.
- o Loading System: Hydraulic truck dumper -- inclined conveyor -- horizontal conveyor -- mechanical plow lowered onto the belt -- material scraped into the LB tank.
- o Unloading System: Suction pumps at each of 8 tank outlets pump the effluent slurry out of the LB.

Cost Reporting

- o LB tank costs: installed capital costs only
- o Material handling costs: installed capital costs and electricity
- o Figure LB-1: Tank Shell Cost Curves
- o Table LB-1: Tank Shell Cost Equations
- o Table LB-2: Tank Shell Costs
- o Table LB-3: Material Handling Cost Equations
- o Table LB-4: Material Handling Costs
- o Figure LB-2: Total Tank and Material Handling Capital Cost Curves
- o Table LB-5: Total Tank and Material Handling Capital Cost Equations
- o Table LB-6: Total Tank and Material Handling Capital Costs
- o Figure LB-3: Total Tank and Material Handling Operating and Maintenance Cost Curves
- o Table LB-7: Total Tank and Material Handling Operating and Maintenance Cost Equations
- o Table LB-8: Total Tank and Material Handling Operating and Maintenance Costs

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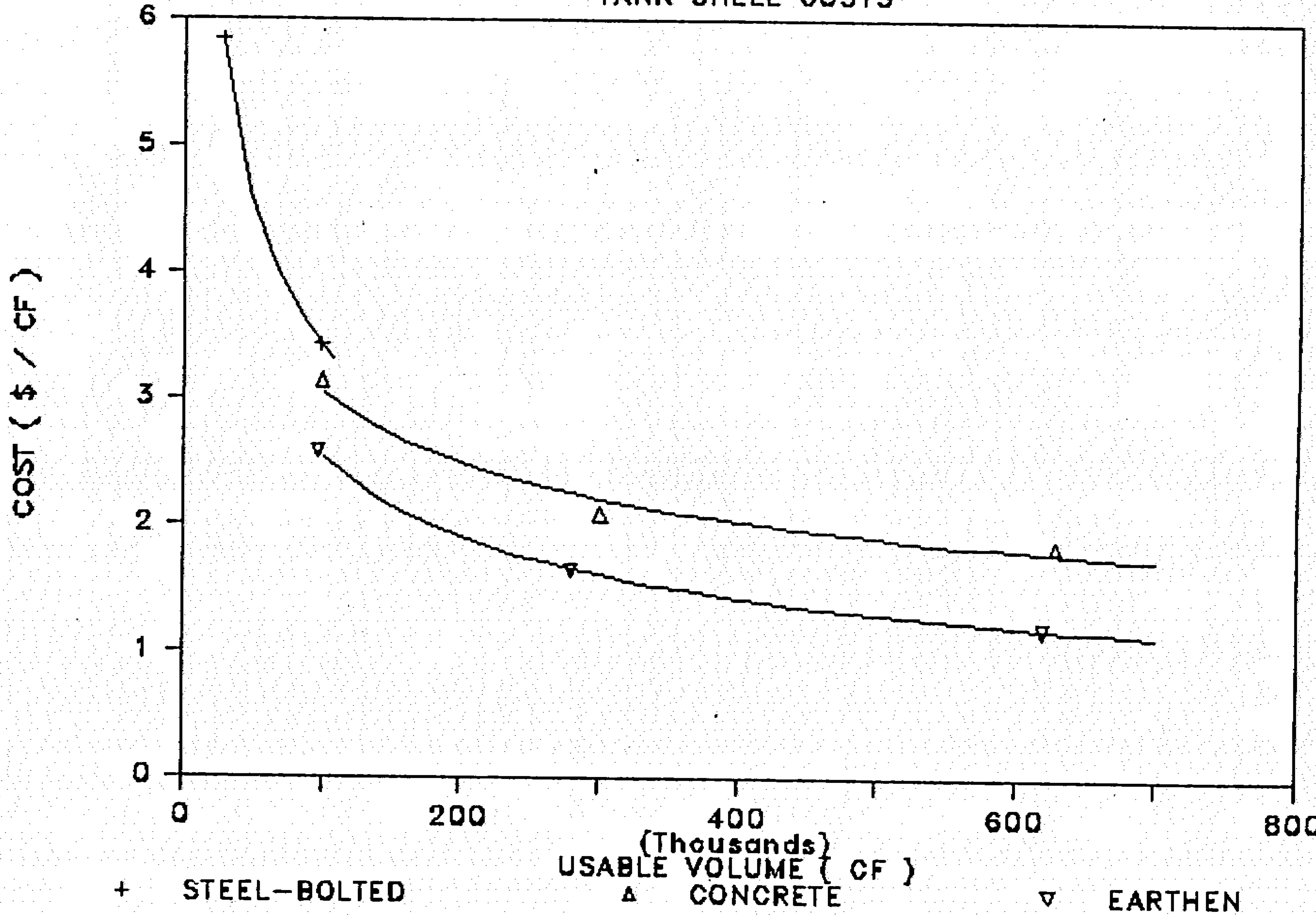
*See section 4.0 for insulation costs

FIGURE LB-1 LEACH-BED REACTOR

TANK SHELL COSTS



ECH4

TABLE LB-1

TANK SHELL COST EQUATIONS
LEACH-BED REACTOR

Steel-Bolted

$$\$/Ft^3 = 364\ V_u^{\ 0.406}$$

S.E. = 0

$\Sigma\ D^2 = 0$

Concrete

$$\$/Ft^3 = 90.2\ V_u^{\ -0.294}$$

S.E. = 0.159

$\Sigma\ D^2 = 0.025$

Earthen

$$\$/Ft^3 = 309\ V_u^{\ -0.417}$$

S.E. = 0.0056

$\Sigma\ D^2 = 0.312 \times 10^{-4}$

VARIABLES:

V_u = Usable volume (ft^3)

S.E. = Standard error

$\Sigma\ D^2$ = Sum of squares of deviations

TABLE LB-2

TANK SHELL COSTS
LEACH-BED REACTOR

	Dimensions (Ft)	Construction	Usable Vol. (Ft^3)[1]	Cost/Vol ($/$Ft^3$)	Refer
1S	42D x 28H	Steel Bolted	26,323	5.850	V, M,
2S	81D x 28H	Steel Bolted	97,907	3.433	V, M, R
1C	80D x 28.5H	Concrete	98,018	3.146	V, M, R
2C	135D x 30H	Concrete	300,591	2.085	V, M, R
3C	200D x 29H	Concrete	628,318	1.844	V, M, R
1E	107.25D x 29H	Earthen	95,832	2.587	V, M, R
2E	163D x 29H	Earthen	279,953	1.648	V, M, R
3E	228.3D x 29H	Earthen	619,777	1.188	V, M, R

(1) Usable Volume = Total - (4 Ft. Rock/Sand Filter + 5 Ft. Freeboard)

ECH6

Insulation costs are shown as a function of insulation thickness.

Insulation Material

- o Sprayed-on polyurethane foam

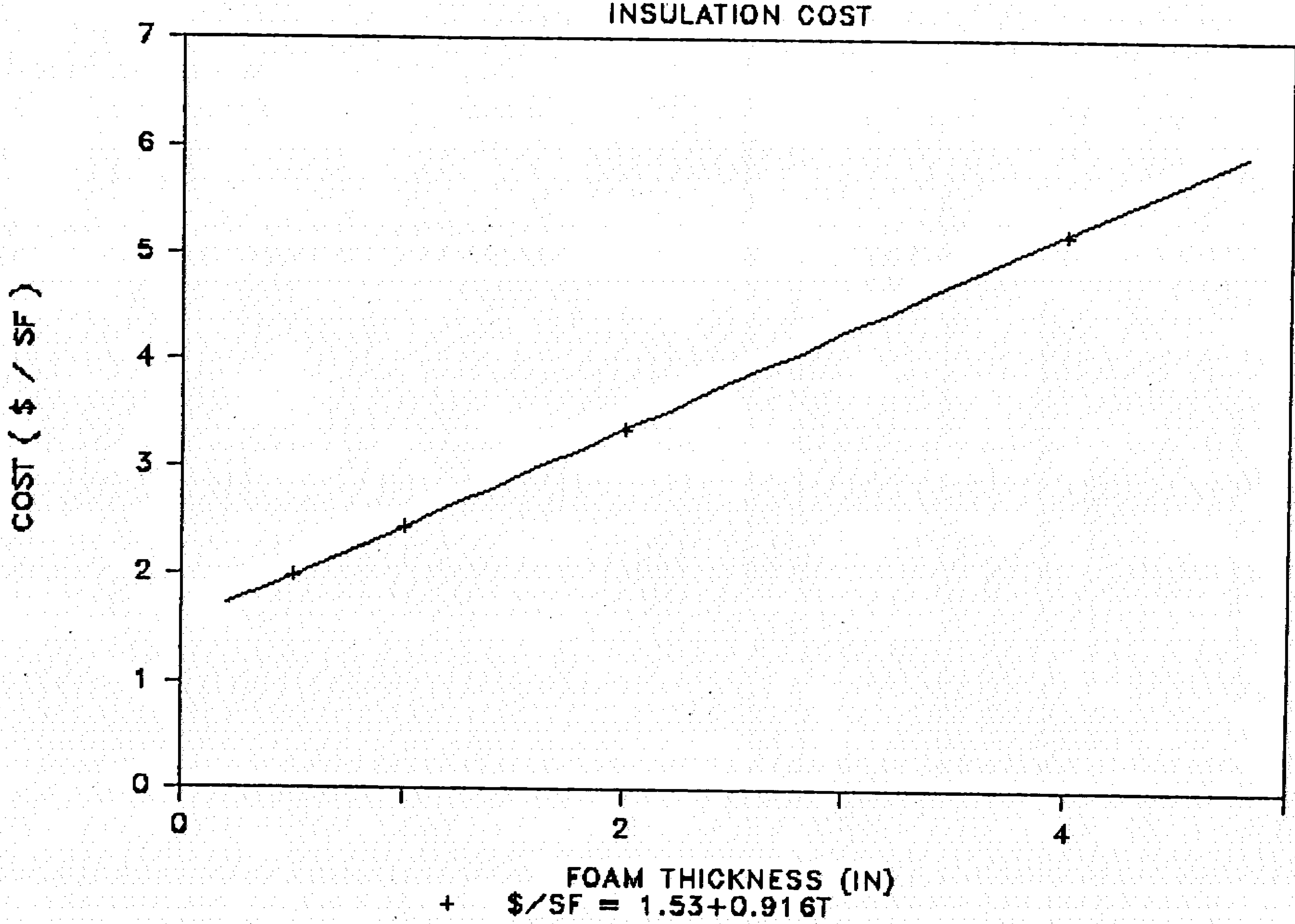
Costs Include

- o Tank surface preparation
- o Surface priming (for proper adhesion)
- o Foam application
- o Exterior protection (for weather and UV light)

Cost Reporting

- o Figure I-1: Cost Curve for Insulation
- o Table I-1: Cost Curve Equation and Costs for Insulation

FIGURE I-1 PACKED BED REACTOR



ECH8

TABLE I-1

INSULATION COSTS
PACKED-BED REACTOR

Foam Thickness (inches)	Cost/Surface Area ($\$/Ft^2$)	Reference
.5	1.984	V
1	2.442	V
2	3.358	V
4	5.189	V

Cost Equation

$\$/Ft^2 = 1.53 + 0.916\ T$ $\quad$ $S.E. = .247 \times 10^{-3}$

$\Sigma\ D^2 = .122 \times 10^{-6}$

VARIABLES:

T = Foam thickness (inches)

S.E. = Standard error

$\Sigma\ D^2$ = Sum of squares of deviations

ECH9

8.1 GAS CLEAN-UP

Unit costs for gas clean-up equipment are shown as a function of biogas input (SCF/day).

Gas Clean-Up System - Monsanto Process

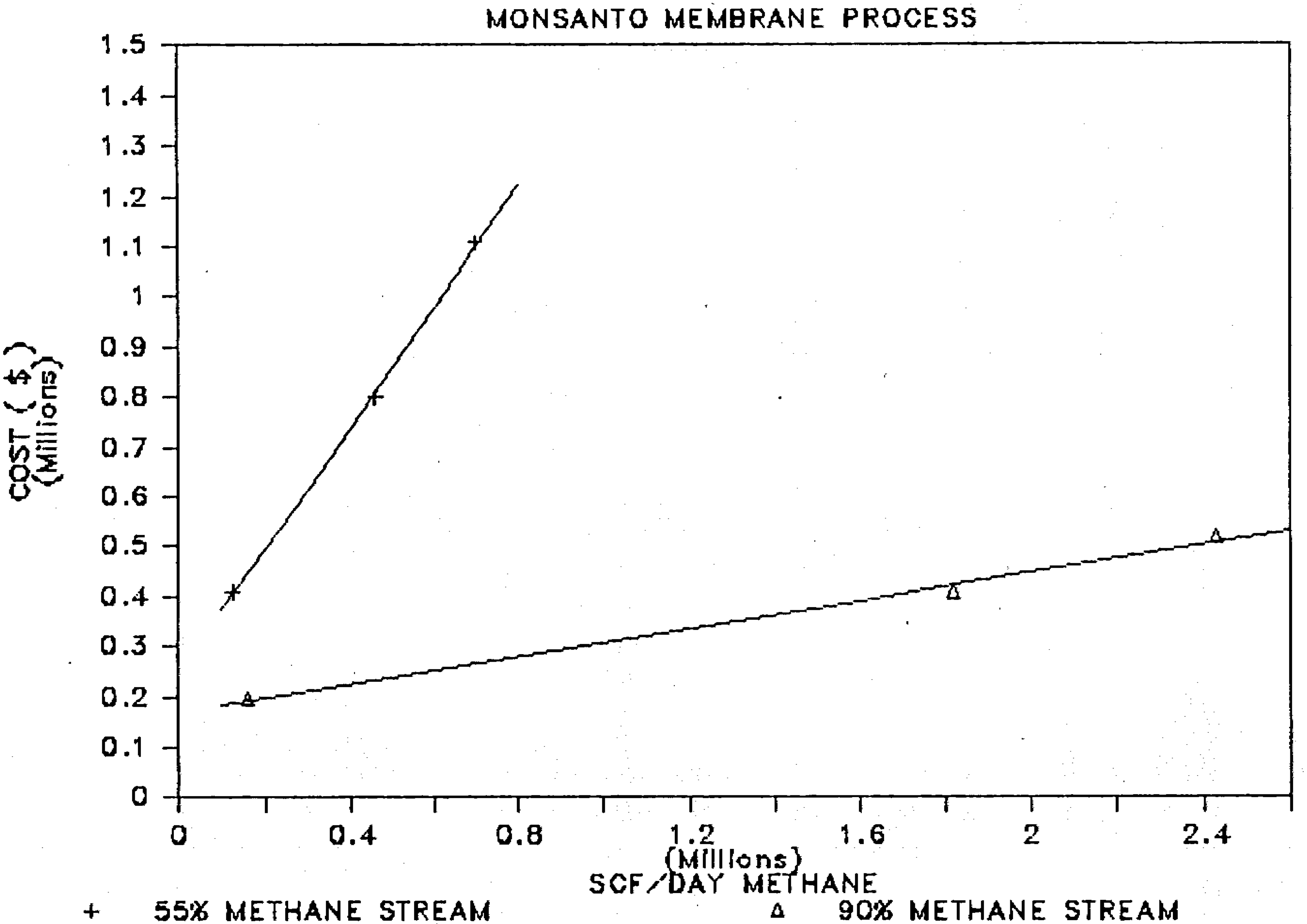
- Selected for costing as the lowest cost technology when compared to Binax or Kryosol gas upgrading processes
- Monsanto technology : gas separation by membrane permeation
- Cost comparisons are included in backup material.

Cost Reporting

- Figure GC1-1: Cost Curves for Gas Clean-up
- Table GC1-1: Cost Curve Equations for Gas Clean-up
- Table GC1-2: Costs for Gas Clean-up

ECH1

FIGURE GCL-1 GAS CLEANUP CAPITAL COSTS



ECH11

TABLE GCL-1

GAS CLEANUP CAPITAL COSTS

SCF/day gas output	% CH_4 in Stream	Cost ($)	Reference
124,000	55	408,605	V
460,000	55	798,794	V
700,000	55	1,108,413	V
160,000	90	195,670	V
1,820,000	90	408,605	V
2,430,000	90	517,950	V

TABLE GCL-2

GAS CLEANUP CAPITAL COST EQUATIONS

55% CH_4: $\$ = 253416 + 1.21150F$

90% CH_4: $\$ = 169613 + 0.13909F$

F = flowrate (SCF/day) of biogas

ECH1

8.2 GAS COMPRESSION

Unit costs for compression as shown as a function of upgraded gas flow (SFC/day). Both capital and operating costs are shown for gas compression after cleanup from 400 to 900 psia.

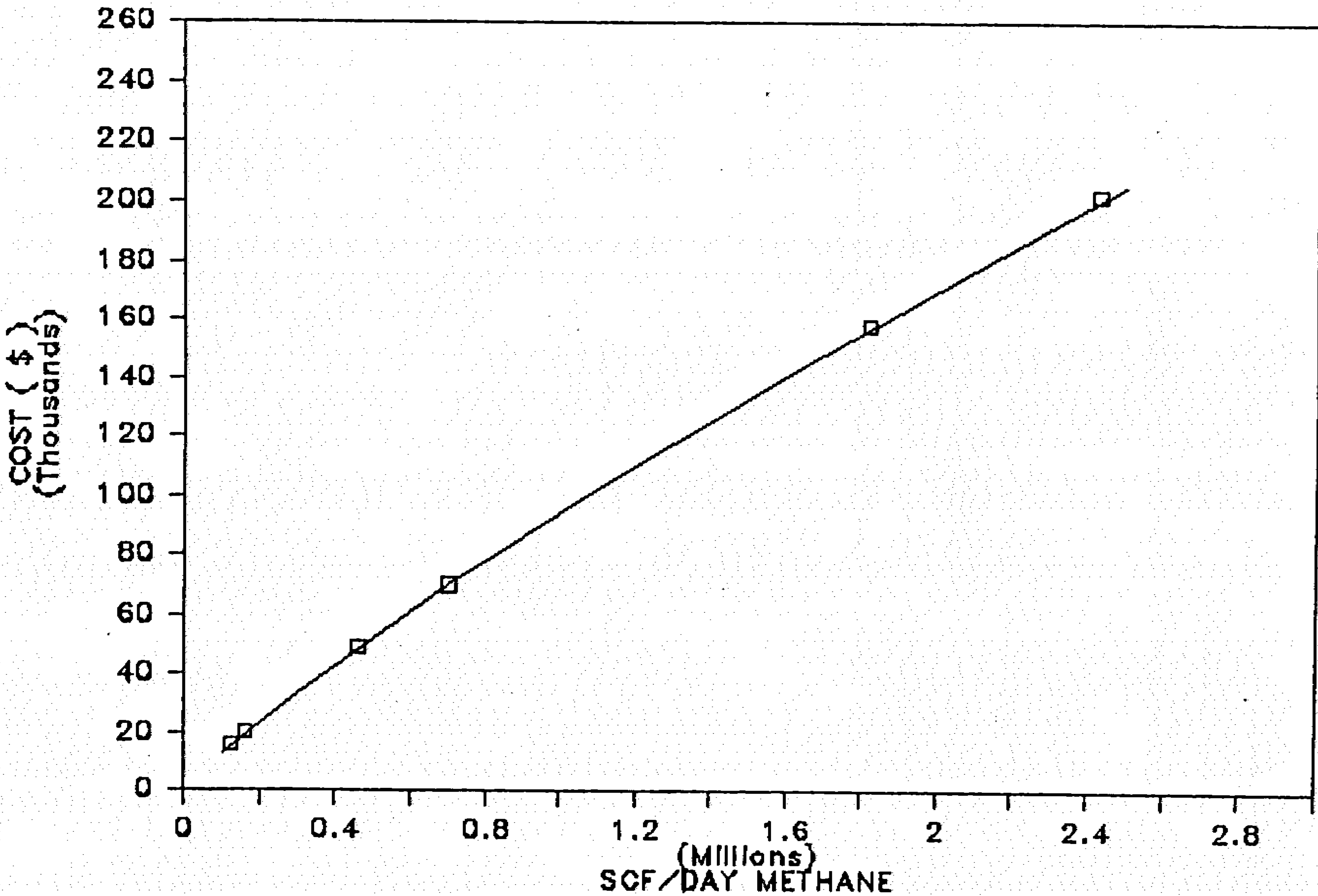
Compression

- o 400 - 900 psia
- o Compressed for injection into pipeline

Cost Reporting

- o Figure GCO-1: Capital Cost Curve for Compression
- o Table GCO-1: Capital Cost Equation and Costs for Compression
- o Figure GCO-2: Operating Cost Curve for Compression
- o Table GCO-2: Operating Cost Equation and Costs for Compression

FIGURE GCO-1 COMPRESSION CAPITAL COSTS



ECH17

TABLE GCO-1

GAS COMPRESSION CAPITAL COSTS

SCF/Day Methane	Cost ($)	Reference
124,000	16,075	V
160,000	19,968	V
460,000	48,962	V
700,000	70,058	V
1,820,000	157,871	V
2,430,000	201,876	V

Cost Equation:

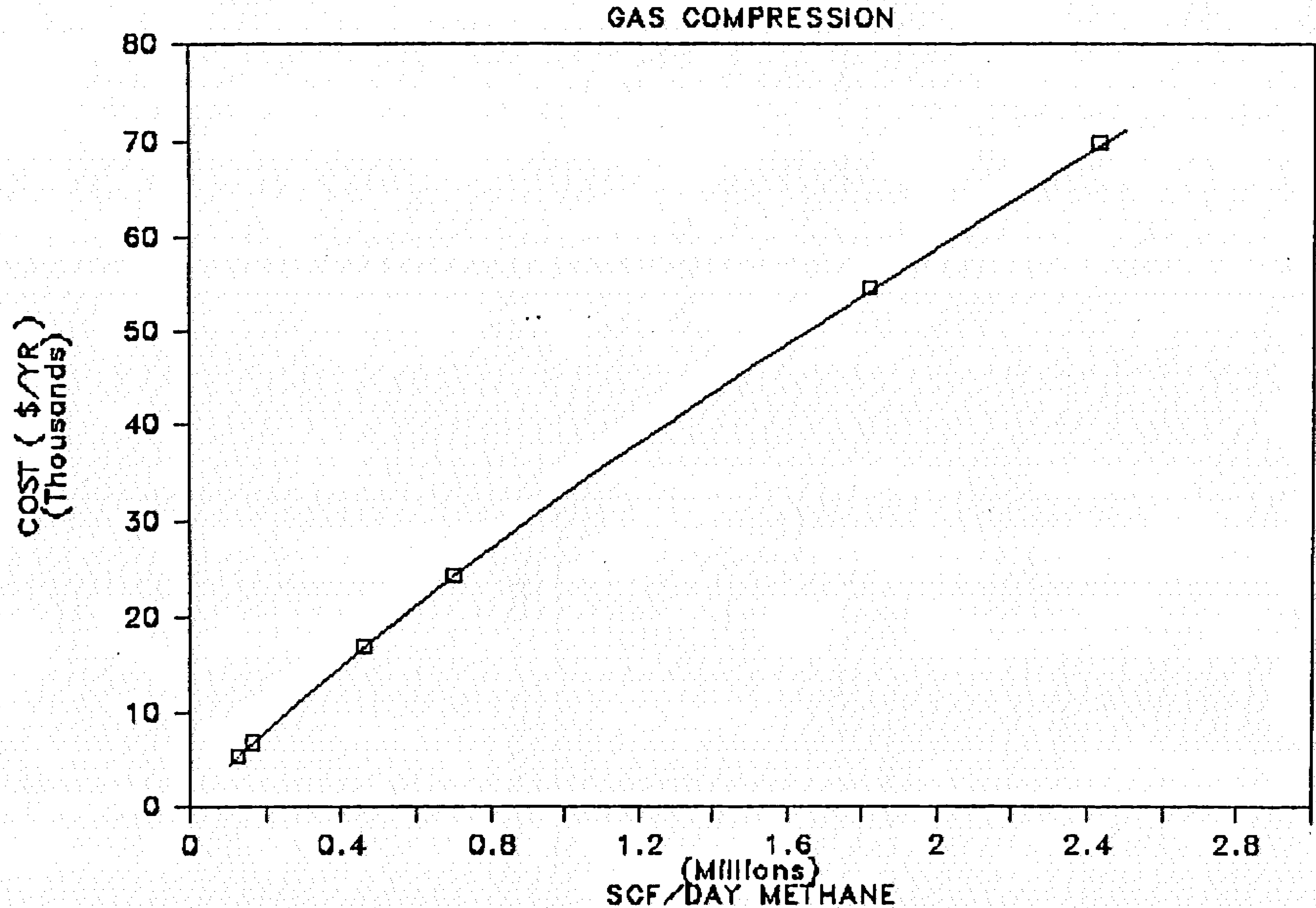
$$\$ = 0.7488 F^{0.850}$$

$S.E. = 30.429$

$\Sigma D^2 = 3703.6$

F = Methane flowrate (SCF/day)

FIGURE GCO-2 OPERATING COSTS



ECHI

TABLE GCO-2

GAS COMPRESSION OPERATING COSTS

SCF/Day Methane	Cost ($/Yr)
124,000	5,556
160,000	6,902
460,000	16,930
700,000	24,215
1,820,000	54,567
2,430,000	69,777

Cost Equation:

$$\$/Yr = 0.2589\ F^{0.850}$$

S.E. = 6.8565

$\Sigma D^2 = 188.05$

F = Methane flowrate (SCF/day)